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Benefits of Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production

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Abstract

The key objective for this project was to evaluate the potential to achieve substantial reductions in the production cost of H₂-rich syngas via coal gasification with near-zero emissions due to the cumulative and synergistic benefits realized when multiple advanced technologies are integrated into the overall conversion process. In this project, Aerojet Rocketdyne's (AR's) advanced gasification technology (currently being offered as R-GAS™) and RTI International's (RTI's) advanced warm syngas cleanup technologies were evaluated via a number of comparative techno-economic case studies.

AR's advanced gasification technology consists of a dry solids pump and a compact gasifier system. Based on the unique design of this gasifier, it has been shown to reduce the capital cost of the gasification block by between 40 and 50%. At the start of this project, actual experimental work had been demonstrated through pilot plant systems for both the gasifier and dry solids pump.

RTI's advanced warm syngas cleanup technologies consist primarily of RTI's Warm Gas Desulfurization Process (WDP) technology, which effectively allows decoupling of the sulfur and CO₂ removal allowing for more flexibility in the selection of the CO₂ removal technology, plus associated advanced technologies for direct sulfur recovery and water gas shift (WGS). WDP has been demonstrated at pre-commercial scale using an activated amine carbon dioxide recovery process which would not have been possible if a majority of the sulfur had not been removed from the syngas by WDP. This pre-commercial demonstration of RTI's advanced warm syngas cleanup system was conducted in parallel to the activities on this project. The technical data and cost information from this pre-commercial demonstration were extensively used in this project during the techno-economic analysis.

With this project, both of RTI's advanced WGS technologies were investigated. Because RT's advanced fixed-bed WGS (AFWGS) process was successfully implemented in the WDP pre-commercial demonstration test mentioned above, this technology was used as part of RTI's advanced warm syngas technology package for the techno-economic analyses for this project. RTI's advanced transport-reactor-based WGS (ATWGS) process was still conceptual at the start of this project, but one of the tasks for this project was to evaluate the technical feasibility of this technology.

In each of the three application-based comparison studies conducted as part of this project, the reference case was based on an existing Department of Energy National Energy Technology Laboratory (DOE/NETL) system study. Each of these reference cases used existing commercial technology and the system resulted in > 90% carbon capture. In the comparison studies for the use of the hydrogen-rich syngas generated in either an Integrated Gasification Combined Cycle (IGCC) or a Coal-to-Methanol (CTM) plant, the comparison cases consisted of the reference case, a case with the integration of each individual advanced technology (either AR or RTI), and finally a case with the integration of all the advanced technologies (AR and RTI combined). In the Coal-to-Liquids (CTL) comparison study, the comparison study consisted of only three cases, which included a reference case, a case with just RTI's advanced syngas cleaning technology, and a case with AR's and RTI's advanced technologies.

The results from these comparison studies showed that the integration of the advanced technologies did result in substantial benefits, and by far the greatest benefits were achieved for cases integrating all the advanced technologies. For the IGCC study, the fully integrated case resulted in a 1.4% net efficiency improvement, an 18% reduction in capital cost per kW of capacity, a 12% reduction in the operating cost per kWh, and a 75–79% reduction in sulfur emissions. For the CTM case, the fully integrated plant resulted in a 22% reduction in capital cost, a 13% reduction in operating costs, a > 99% net reduction in sulfur emissions, and a reduction of 13–15% in CO₂ emissions. Because the capital cost represents over 60% of the methanol Required Selling Price (RSP), the significant reduction in the capital cost for the advanced technology case resulted in an 18% reduction in methanol RSP. For the CTL case, the fully integrated plant resulted in a 16% reduction in capital cost, which represented a 13% reduction in diesel RSP. Finally, the technical feasibility analysis of RTI's ATWGS process demonstrated that a fluid-bed catalyst with sufficient attrition resistance and WGS activity could be made and that the process achieved about a 24% reduction in capital cost compared to a conventional fixed-bed commercial process.

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1. Executive Summary

In this project, the key accomplishment has been the thorough and practical evaluation of several emerging technologies for reducing the production cost of power and chemicals from coal, one of the United States' most abundant resources, with an equally important goal of also achieving emission levels that match or even exceed similar processes used for the conversion of natural gas or oil-derived hydrocarbons. A key criterion for the selection of these emerging technologies was a commercialization horizon that was between 2 and 5 years to ensure that any benefits identified would be available for commercial application immediately or at least very soon. A second criterion was that these technologies were in the pilot plant or demonstration phase of their development, which would ensure that existing data on operational performance and cost would reflect testing at typical commercial operating conditions. The availability of this actual data significantly increased the quality and accuracy of the evaluations completed.

The specific emerging technologies selected included Aerojet Rocketdyne's (AR's) advanced gasification technology and RTI International's (RTI's) advanced warm syngas cleanup and conditioning technologies. AR's advanced gasification technology includes a dry solids pump that effectively pumps coal particles from atmospheric to high pressure and an extremely compact gasification system that effectively reduces the volume of the gasifier itself by up to 90%. At the start of this project, this technology was undergoing pilot plant testing of these components at several research facilities. Currently, this technology package has been acquired by the Gas Technology Institute (GTI) and is being offered under the trade name of R-GAS™.

The breakdown of RTI's advanced warm syngas cleanup and conditioning system includes their Warm Gas Desulfurization Process (WDP), Direct Sulfur Recovery Process (DSRP), and an Advanced Fixed-bed Water Gas Shift process (AFWGS). A key difference between WDP and conventional Acid Gas Removal (AGR) processes is that WDP removes both H₂S and COS, but not CO₂. This effectively decouples sulfur and CO₂ removal which leads to opportunities for simultaneous improvements in cost and efficiency of removing sulfur and CO₂. The actual demonstration of this was done with the pre-commercial demonstration of RTI's advanced cleanup and conditioning system coupled with BASF's aMDEA™ (activated amine process) in a pre-commercial demonstration plant that processed a 50 MW_e equivalent stream of syngas from Tampa Electric's Integrated Gasification Combined Cycle (IGCC) facilities at their Polk Power Station in Tampa, FL. With over 3,500 hours of successful operation of this pre-commercial demonstration system as a reference, RTI has partnered with Casale SA to offer this technology commercially.

The AFWGS process, which was part of RTI's 50 MW_e pre-commercial demonstration plant, demonstrated a novel reactor configuration based on available commercial water gas shift (WGS) catalyst that significantly reduced steam consumption in the WGS process that was required for the conversion of carbon monoxide into hydrogen and CO₂. By reducing this parasitic steam consumption, AFWGS effectively enables more steam to be put to beneficial use generating additional power in the steam turbine, which increases overall efficiency. Encouraged by the success of AFWGS, RTI envisioned a new Advanced Transport Water Gas Shift (ATWGS) process that leverages existing technical knowhow from Commercial Fluid Catalytic Cracker (FCC) processes to improve the thermal efficiency even more than AFWGS.

As this ATWGS technology was less mature than AFWGS, the work in this project was confined to assessing the technical feasibility of this technology, effectively setting the stage for rapid development of this technology within the next 5 to 10 years. A key component of this technology would be the fluidizable WGS catalyst. Existing commercial catalysts have performance that has been optimized over several decades, but they are only available in formulations that are suitable for fixed-bed reactors. In addition to being composed of roughly spherical particles within a set size range, a suitable fluidizable catalyst must also have the strength or attrition resistance to withstand the abuse associated with moving around the transport reactor system as well as having catalytic activity equivalent to commercial fixed-

bed catalysts. All of the small set of catalysts prepared as part of this project for the three different commercial applications of the WGS process had attrition resistance better than the standard used for commercial FCC catalysts. From catalytic testing, the catalyst formulation for high temperature sweet water gas shift also had catalytic activity comparable to a commercial fixed-bed water gas shift catalyst. This demonstrated a near-ready catalyst formulation for scaling up to pilot plant testing of ATWGS. The economic evaluation of the ATWGS process showed a 24% reduction in capital cost compared to a conventional fixed-bed WGS process.

The evaluation of the more mature emerging technologies consisted of a series of case studies for three coal conversion applications in which the performance of a plant using the best available commercial technologies were compared with plants incorporating the advanced emerging technologies. Techno-economic analyses were used to establish the performance of plants incorporating the emerging technologies. The plants based on the best available commercial technologies were obtained from techno-economic analyses completed by DOE/NETL in published studies. In addition, all the techno-economic analyses completed for the plants with the emerging technologies followed the specific and detailed guidelines DOE/NETL established for their own energy studies so as to ensure the most accurate possible comparison of technologies.

The selection of the coal conversion applications focused on those that would be of most benefit/impact for the public. The first of these applications was the generation of electricity using an IGCC with > 90% carbon capture. The second was the production of methanol, which is one of the fundamental raw materials used throughout the chemical industry and can also be used as a fuel. A unique aspect of the plants studied in this case study was the capture of sufficient CO₂ byproduct to enable >90% carbon capture. The final application was for Fischer-Tropsch (FT) synthesis of naphtha and diesel products. This application supports the large consumer-driven transportation fuels market. As with the power and methanol applications, the plants evaluated in this FT case study captured the CO₂ byproduct to achieve >90% carbon capture.

In an effort to optimize our understanding of the benefits and fully understand any synergistic benefits associated with integrating the emerging technologies, multiple advanced plant configurations were evaluated in the case study for each of the three applications. For the IGCC with > 90% carbon capture application, the case study evaluated three advanced plant configurations in addition to the reference plant based on best commercial technologies. Two of these advanced plants incorporated each of the emerging technology packages (AR or RTI) by themselves. The last advanced plant integrated both emerging technology packages (AR and RTI). This same selection of advanced plant configurations was also used for the Coal-to-Methanol (CTM) application. The Coal to Fischer-Tropsch Liquids (CFTL) application only included two advanced plant configurations for comparison with the reference case. One of these was the advanced plant with the integration of both emerging technology packages. The other advanced plant included only the RTI advanced syngas cleanup and conditioning technology package. A key differentiator for the advanced plants in the case studies conducted under this project was that the integration process was led by the actual technology developers, who would have the best insights to overcoming any challenges that arise when integrating their processes. Beyond their technical insights relating to their technologies, these developers also had access to the latest pilot plant and demonstration plant data and costing, which raised the level of accuracy associated with the techno-economic analyses for these emerging technologies.

Although in all three of the case studies the plant configurations with a single emerging technology did result in substantial benefits, the best overall benefits were consistently observed for the plants with the integration of both emerging technology packages. Because the objective of this project was to achieve optimal benefits in terms of cost and emissions reduction, the focus in this executive summary is on discussion of the advanced plants integrating AR's and RTI's advanced technologies. However, these case studies do provide a wealth of information on the individual emerging technologies and we encourage the reader to dig deeper in the report to learn more about these technologies and their individual benefits.

By design, one of the key performance metrics that was evaluated for each plant in each of the cases studies was the production cost or required selling price (RSP) for the power, methanol, or FT liquids (diesel and naphtha) products. The secondary byproducts, which included electricity and CO₂ were assigned a value as per DOE/NETL guidelines, but the impact of the selected value on the RSP was evaluated in the sensitivity studies. The integrated advanced plants reduced RSP and capital and operating costs in all three case studies as can be seen in **Table E-1** which summarizes the overall benefits.

Table E-1. Summary of benefits identified in case studies

	IGCC	CTM	CFTL
Product cost reductions	18 mills/kWh	\$77–93/ton	\$17–21/bbl ¹
Capital cost reduction	18%	23%	15%
Operating cost reduction	12%	13%	3.6%
Efficiency Improvement	1.4%		
Surplus power generated		28 MW ²	41%
Emission reductions			
Sulfur	75–79%	99%	
CO ₂ emissions		13–15%	4 lb CO ₂ /bbl ³

¹ This cost reduction applies for both the diesel and naphtha products produced.

² The reference case consumed approximately 1 MW of external power.

³ The CFTL cases inherently remove the CO₂ to optimize the FT process, but the advanced technologies manage to reduce net CO₂ emission per unit of product produced.

As would be expected, capital cost was the largest single contributor to the RSP for all three case studies typically contributing > 50% of the RSP. Both AR's and RTI's technologies contributed to overall plant capital cost reductions of between 15% and 23% depending on the product application. Both these emerging technologies also resulted in a significant reduction in operating costs. The reduction in the operating cost for the CFTL case was lower than that for the IGCC and CTM cases for two primary reasons. The first is that the main contributor to the operating cost was the replacement cost for the FT catalyst, which neither emerging technology had any influence over. The second was that the coal fuel for the CFTL case was an Illinois #6 coal, which has a high chlorine content. The removal of this chlorine for environmental as well as process protection reasons increased operating expenses. For most available coals, which generally do not have this high chlorine issue, a higher reduction in the operating expense would also be achieved for the CFTL case.

The results in Table E-1 also show that the emerging technologies resulted in significant efficiency improvements that were recovered as additional power generation. The challenge was that for the CTM and CFTL cases, this power was an inherent byproduct, and as such, market forces would typically require a lower selling price. Although the sensitivity analysis did show that the selling price of the electricity had essentially no impact on the RSP, this is bad because a significant byproduct and benefit of the emerging technologies is being undervalued. Alternative means of utilizing this byproduct within the plant should be explored to maximize its value.

The final elements in Table E-1 show how the emerging technologies, and in particular RTI's warm syngas technology, reduced emissions of sulfur and CO₂. The fact that WDP specifically targets the removal of both H₂S and COS, but not CO₂ enables the decoupling of sulfur and CO₂ removal that is not possible in most conventional AGR processes. The consequence of this decoupling is that significantly more sulfur and CO₂ can be removed during syngas cleaning. This is best illustrated by the CTM case study. The IGCC case study does show the improvement in sulfur, but there was no improvement in CO₂ emissions. This is because removal of the CO₂ significantly reduces power generation in the gas turbine since the mass of the CO₂ has been removed from the syngas and because of the auxiliary power load

required to compress the CO₂ for sequestration. For the IGCC case, the optimal power generation performance is achieved when the 90% carbon capture is just met. For the CFTL case, stringent contaminant removal is required for the FT processes. Under these constraints, the resulting emissions are already very low. However, RTI's technologies are still able to add benefits. For the CO₂, this is a reduction of 4 lb of CO₂/bbl. For sulfur, the benefit of RTI's syngas cleaning technologies is not seen in the emissions, where RTI's technologies achieve a syngas with contaminant concentrations comparable to those generated by a conventional Rectisol[®] plant, but is seen in the significant reduction in capital cost for syngas cleanup, which is reflected in the reductions in RSP and overall capital costs.

Substantial benefits from the integration of AR's advanced gasification technology and RTI's advanced warm syngas cleaning and conditioning technologies, which include capital cost reductions, lower sulfur and CO₂ emissions, and improved energy efficiency were consistently observed in the three case studies focused on power, methanol and FT liquids production. The fact that these benefits were still observed in sensitivity studies even when the capital cost of these emerging technologies were increased by 30% confirms the robustness of these benefits. To realize these benefits, market forces will drive the implementation of these emerging technologies. The result is that as these technologies begin to penetrate the market, these technologies will be favored over existing technologies and result in a significant reduction in emissions associated with coal utilization.

2. Introduction

The fundamental challenge for gasification for power and coal-to-liquid (CTL) production is to achieve production costs for syngas that ensure the overall conversion process is cost-competitive with other conversion technologies and alternative feedstocks such as natural gas. Previous studies by the U.S. Department of Energy (DOE) and others have indicated that this challenge cannot be addressed by improving only one step in the overall gasification process. The overall objective of this project is to assess the potential to achieve a substantial reduction in the production cost of H₂-rich syngas via gasification with near-zero emissions due to the cumulative and synergistic improvements realized when multiple advanced technologies are incorporated into the overall conversion process. Our selection of advanced technologies includes several of the most advanced and potentially cost-disruptive technologies under current development: the compact gasification system (CGE) and dry solids pump (DSP) being developed by Aerojet Rocketdyne (AR, and currently offered by the Gas Technology Institute as R-GAS™), warm syngas cleanup technologies and a novel water gas shift H₂ enrichment process being developed by RTI International (RTI), and an activated amine carbon dioxide recovery process (AACRP) already commercially developed and offered by a number of vendors including BASF, UOP, and Shell. The AACRP offers substantial benefits for carbon dioxide capture, but has never before been applied to syngas applications due to constraints that are eliminated when integrated with RTI's technology.

Through previous individual techno-economic studies, AR's advanced gasification with dry feed and RTI's warm syngas cleanup technologies have shown the potential to enable substantial reductions in cost and significant improvements in efficiency to the overall gasification process for power generation. These benefits have now been substantiated through extensive pilot plant or near-commercial demonstration testing. Beyond the cumulative benefits, additional synergistic benefits are likely to be realized from combining the proposed technologies. Such synergistic benefits would most likely result from reductions in raw material usage (i.e., reduced production costs) and equipment size (i.e., reduced capital costs) due to reduced throughputs achieved from efficiency improvements associated with technology integration.

By engaging the key technical experts from AR and RTI, and utilizing the best available data on costs and performance from recent pilot plant and near-commercial demonstrations of their technology, optimal integration of AR's and RTI's technologies could be achieved that are specific for ultra-clean power generation or CTL production with near-zero emissions with the latest and best available technical and cost data. With Nexant leveraging their recognized engineering and design capabilities to perform the system studies, these system studies offered the potential to demonstrate a disruptive technology change associated with the cumulative integration of multiple advanced technologies that could fundamentally redefine the market.

3. Advanced Technology Descriptions

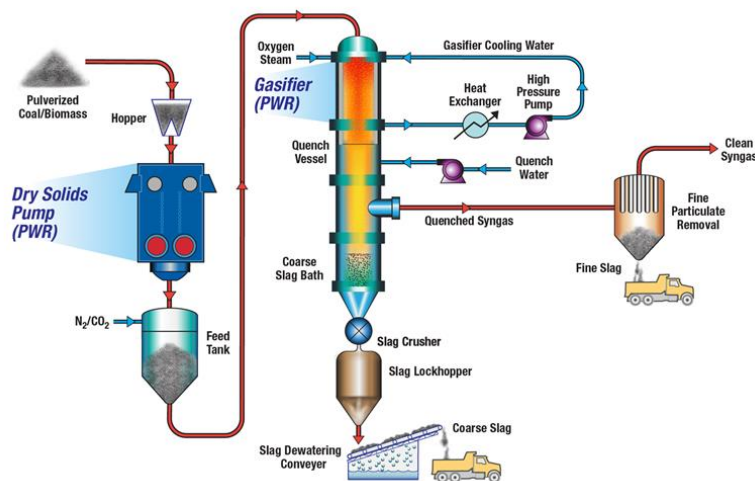


Figure 1. Schematic of AR advanced gasification technologies

Descriptions of the specific process technologies included in these coal to H₂-rich syngas systems are provided in **Table 1** along with the current-state-of-the-art competing technologies. **Table 2** provides a summary of the benefits associated with the proposed technologies along with the techno-economic analyses and supporting experimental data. The information for AACRP integrated with RTI's warm gas desulfurization process (WDP) and/or the transport reactor-based water gas shift H₂ enrichment processes in **Table 3** contains the expectations

anticipated at the start of this project based on results for similar techno-economic analyses or projections from existing techno-economic analyses.

3.1 Aerojet Rocketdyne's Advanced Gasification Technology

A simplified schematic of AR's gasification technology is shown in **Figure 1**. The key components of AR's technology include the DSP, feed splitter, high mixing efficiency injectors, advanced gasification internals cooling and rapid spray quench designs. AR has performed testing in an 18 ton per day (tpd) pilot plant gasifier at the Gas Technology Institute (GTI) facility in Des Plaines, IL. **Figure 2** shows a photograph of this pilot plant gasifier. During this testing, AR accumulated 764 hours of hot-fire operation and completed 53 mass and energy balance tests on Illinois #6 coal, oil sands petcoke, petcoke from a commercial refinery, and Alberta sub-bituminous coal. **Figure 3** shows the mass balance results for this testing. These results from this pilot plant testing have been used to anchor a computational fluid dynamics (CFD) model of the gasifier to facilitate scale up of the gasifier design for commercial application. Similar testing of AR's DSP system has been used to develop the fundamental components and to investigate its operation. **Figure 4** shows the DSP testing systems and key results collected during testing. The results have enabled development of a 400 tpd system shown in **Figure 5** currently installed at the Energy and Environmental Research Center (EERC) in Grand Forks, ND, where testing is in progress. Based on this activity, AR anticipates that their technologies should be in commercial demonstration by 2018.

Table 1. Description of proposed processing technologies

Proposed technology	Description	Current-state-of-the-art competing technology
AR's dry solid pump (DSP)	Continuous pressurization of the coal feed stream from ambient up to 1,200 psig.	• Dry lock hopper feed systems • Slurry-based feed systems
Compact gasification system (CGE) currently being offered by GTI as R-GAS™	The compact gasification system consists of AR's feed splitter, high mixing efficiency injectors, advanced gasifier internals cooling, and rapid spray quench systems.	• Entrained-bed gasifier • Injectors • Refractory lining and/or water cooled lining • Deluge quench systems
Warm Gas Desulfurization Process (WDP)	This consists of a warm desulfurization process that is capable of simultaneous removal of H ₂ S and COS to < 10 ppmv using a regenerable fluidizable sorbent in a dual transport reactor. In the adsorber, H ₂ S and COS are removed and in the regenerator the sulfur is extracted as a SO ₂ byproduct.	• Physical solvent-based systems (Selexol™ or Rectisol®)
Direct Sulfur Recovery Process (DSRP)	A fixed bed catalyst system that partially reduces the SO ₂ byproduct from regeneration with a slipstream of syngas into elemental sulfur at regeneration off gas conditions.	• Claus Plant and Tail Gas Treatment
Advanced Transport Water Gas Shift Process (ATWGS)	A novel transport reactor that incorporates solids cooling technology used extensively in commercial fluid catalytic cracking to provide temperature control for exothermic water gas shift reaction.	• Fixed-bed water gas shift reactors (multiple stages with interstage cooling)
AACRP	Non-selective solvent based CO ₂ and sulfur removal system commercially demonstrated in over 300 operating plants.	• Non-selective amine CO₂ capture processes or Selexol™

Table 2. List of the proposed processing technologies and benefits

Proposed technology	Benefits	Techno-economic studies	Supporting experimental data
AR's dry solid pump (DSP)	32 GWh/y reduction in lock hopper power requirements (3,000 tpd plant) - Demonstrated performance with both coal and coal/biomass mixtures		
Compact gasification system (CGE), currently being offered by GTI as R-GAS™	90% reduction in gasifier volume - Cold gas efficiency improvements 7%–10% versus water slurry feeders 2%–4% versus dry feed systems - Gasifier surface temperatures of 1000°F > 2 year life injectors (< 4 months GE injectors) > 10 year life cooling liner (1–3 years for refractory) - Dry ash recovery eliminates black water collection system and waste water treatment requirements. > 15% lower cost of electricity for IGCC > 25% lower cost for hydrogen	- DOE/NETL (comparison of AR IGCC and Commercial IGCC Performance), 2006 - Advanced IGCC report by Jacobs Engineering for the Canadian Clean Power Coalition Technical Committee, 2011	- Feed splitter 400 tpd has been tested at EERC 18 tpd gasifier was tested between 2009 and 2011 at GTI
Warm Gas Desulfurization Process (WDP)	>3 percentage point increase in HHV thermal efficiency for IGCC 10%–15% reduction in overall capital cost for IGCC system without carbon capture	- ChevronTexaco: RTI HTDP and DSRP Technology Evaluation and Comparison, 2003 - Nexant: Preliminary Feasibility Analysis of RTI Warm Gas Cleanup Technology, 2007 - DOE/NETL: A Pathway Study in Non-carbon Capture Advanced Power Systems R&D using Bituminous Coal, Volume 1, 2008)	0.3 MW_e field test (DE-AC26-99FT40675) 50 MW pre-commercial demonstration (DE-FE-0000489)
Advanced Transport Water Gas Shift Process (ATWGS)	- Reduces capital cost by eliminating multiple reactor stages with interstage cooling (> 45% capex reduction for transport reactor-based vs multiple stage fixed-bed methanation process). - Achieves higher conversion by operating at lower temperatures. - Reduces steam requirements by reducing need for steam to control exothermic reaction.	RTI techno-economic studies for transport reactor-based methanation process for syngas to substitute natural gas (SNG) conversion	- Pilot plant testing demonstration of syngas conversion to SNG at KBR. - Prototype transport reactor testing at RTI for syngas conversion to SNG.

Proposed technology	Benefits	Techno-economic studies	Supporting experimental data
AACRP	• Higher adsorption capacity and kinetics • Lowest specific energy consumption for any standard amine • In combination with WDP (for S and CO₂ capture) • 20–35% capex reduction vs Rectisol® • ~20% capex reduction vs Selexol™	Internal techno-economic assessment by RTI using Nexant's preliminary feasibility study as a basis	BASF began licensing its aMDEA™ technology in 1982. It currently has over 300 reference plants in operation around the world; integrated demo plant with RTI WDP demonstrated.

Benefits of Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production



Figure 2. AR compact gasifier system at GTI

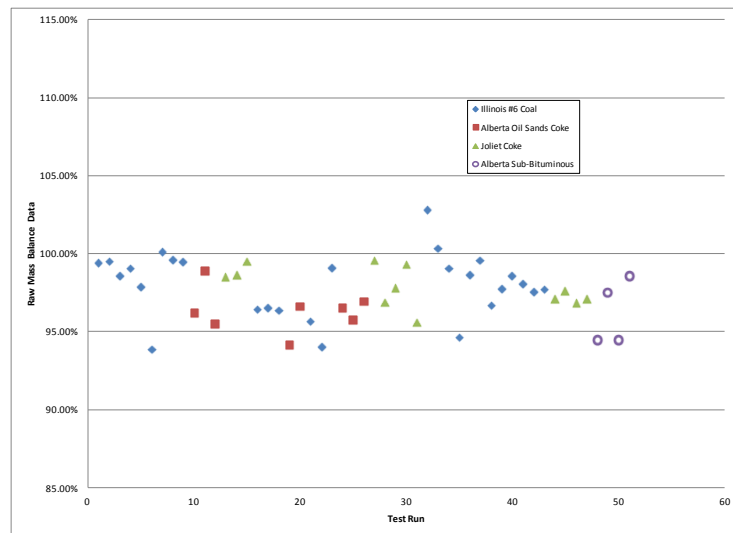


Figure 3. Mass balance data from AR gasification tests at GTI

<u>Flow Tester</u>	<u>Wall Tester</u>	<u>Friction Tester</u>	<u>Permeability Tester</u>
~200-300 psi available for deconsolidating forces	- Track Aspect Ratio - Maximum duct solids pressure	- Outlet duct size for gas seal - Link load stress analysis	- Confirm seal duct length - Leak rate at 1200 psi pressure
			

Figure 4. Component test units for AR dry solids pump

3.2 RTI's Advanced Syngas Cleanup Technologies



Figure 5. 400 TPD AR dry solids pump installed at EERC

construction and operation of a 50 MW_e syngas cleanup system with carbon capture using a 20% slipstream of the total syngas produced at Tampa Electric Company's (TEC's) Polk Power IGCC system. **Figure 8** shows the 3D engineering model of this system and a photograph of the completed construction. Operation of this system began in early 2014 and was completed in April 2016. This demonstration unit is expected to be readily scalable to a commercial-scale plant.

Because of demonstration plant schedule constraints, a commercially available carbon capture process was chosen for the 50 MW_e system at TEC. Selection of

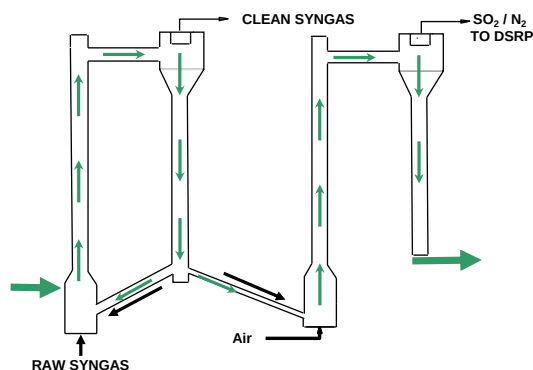


Figure 7. Schematic of the RTI WDP transport reactor system

50% less than the standard Rectisol[®] process—the commercial standard for the high purity requirements necessary for chemical production. Although AACRP technologies are offered by multiple vendors

The key technologies that form RTI's warm syngas cleanup technologies include the WDP and the direct sulfur recovery process (DSRP), (see descriptions in Table 2). These technologies can be effectively integrated to achieve near-zero emissions. Pilot plant testing of RTI's WDP and DSRP were completed with real coal-derived syngas at Eastman Chemical Company's coal-to-chemicals facility in Kingsport, TN.

During this field test at Eastman, over 3,000 hours of operation of the WDP were completed demonstrating greater than 99.9% removal of H₂S and COS. DSRP and warm trace contaminant removal were also successfully demonstrated during this field test with between 100 and 500 hours of operation. **Figure 6** shows the pilot plant systems installed at Eastman. **Figure 7** shows the reactor schematic for this pilot plant WDP. Under a current DOE project (FE-0000489), RTI is actively working on the design,



Figure 6. RTI warm syngas cleanup pilot plant systems at Eastman

this process included evaluation of a number of commercial physical and chemical solvent-based processes. An AACRP offered many distinct advantages especially when integrated with RTI's WDP. Because the AACRP is non-selective for sulfur and CO₂, it is not a standard choice for conventional syngas applications despite its excellent stand-alone performance for CO₂ capture. However, because RTI's WDP removes a majority of the sulfur before this process, the AACRP when integrated with WDP will remove > 98% of the CO₂ and > 99% of the remaining sulfur. The combination of RTI's warm syngas cleanup technologies and AACRP generates a syngas stream with near-zero contaminants concentrations. The most impressive benefit of this combination is that the capex of this technology combination is estimated to be 20 to

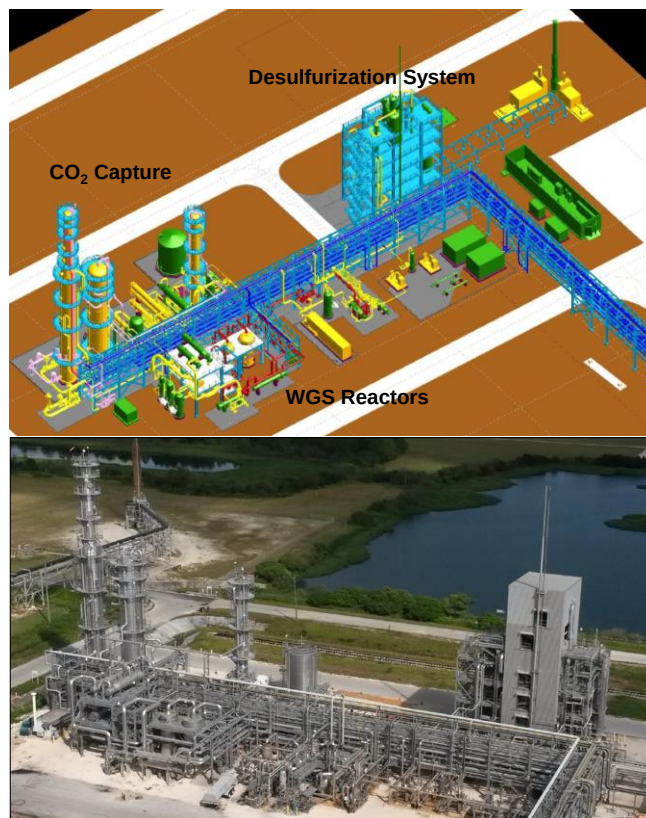


Figure 8. 50 MW pre-commercial unit at Tampa Electric Company a) 3D engineering model and b) photograph of completed system

including BASF, UOP and Shell, RTI selected BASF's aMDEA™ for integration with their pre-commercial demonstration plant at TEC. One of the key selling points was commercially demonstrated cost benefits of BASF's aMDEA™ process, which are based on over 300 operating plants around the world and BASF's process licensing efforts since 1982. RTI has already addressed integration issues of WDP and AACRP during design of the 50 MW_e syngas cleanup system with carbon capture at TEC.

3.3 RTI's Advanced Water Gas Shift Processes

All gasification technologies for coal generate a syngas mixture that has high concentrations of CO. For chemical and power production with carbon capture, this CO must be shifted into H₂ via the water gas shift (WGS) reaction, $\text{CO} + \text{H}_2\text{O} = \text{CO}_2 + \text{H}_2$. The key technical challenge is that the WGS reaction is exothermic with a ΔH of -41 kJ/mole. Consequently, syngas mixtures with higher CO concentrations generate more reaction heat, which increases the reactor temperature and reduces the amount of conversion possible as

equilibrium limitations for exothermic reactions reduce conversion at higher temperatures. Commercial fixed-bed WGS technologies use the addition of steam and multiple reactor stages with interstage cooling to achieve the necessary amount of overall CO shift. Because it is a reactant in the WGS reaction, steam enhances the equilibrium conversion of CO. Steam also has a very high heat capacity allowing the excess steam to provide additional thermal mass to absorb the enthalpy released by the WGS reaction. The consequence of applying these standard WGS technologies to achieve H₂-rich syngas is the increased capex associated with the multiple reactors and decreased thermal efficiency from the increased steam consumption. For high carbon capture requirements, steam consumption by the WGS system increases the parasitic power load significantly.

For the 50 MW_e pre-commercial demonstration plant, the high steam consumption required by a conventional commercial WGS system exceeded the amount of available high pressure steam without significantly derating the total power output of TEC's Integrated Gasification Combined Cycle (IGCC) plant. As a consequence, RTI was forced to develop a modified WGS system that significantly reduced the high pressure steam consumption in the WGS system. This modified WGS system used available commercial WGS catalyst in a novel fixed-bed reactor configuration. Through the pre-commercial demonstration testing completed at TEC, this Advanced Fixed-bed Water Gas Shift (AFWGS) process was successfully operated for over 680 hours while achieving over 90% conversion of the CO in the syngas. Because of its advantages, this AFWGS process was used in all the process system studies completed in this project for power generation with carbon capture and CTL with near-zero emissions.

As part of the original scope of this project, RTI also proposed an Advanced Transport reactor-based WGS (ATWGS) process. This novel ATWGS process, shown in **Figure 9**, leverages commercial technologies used for Fluid Catalytic Cracking (FCC). These technologies include the transport reactor and a solids cooler. In a transport reactor, the reactant gases are mixed with small fluidizable catalyst

particles. The reactant gas carries these catalyst particles while moving through the reactor. The intimate mixing of the catalyst and reactant gas promotes excellent mass and heat transfer. The result is the reactant gases react and any heat generated heats the catalyst particles as well as the gas. At the reactor outlet, the catalyst particles are separated from the product gas, usually with a cyclone, and the catalyst is transferred back to the reactor inlet. During this transfer, the catalyst particles are cooled in a solids cooler enabling the catalyst particles to provide a regenerable thermal mass for temperature control in the WGS reactor. This approach should lead to much higher conversion than can be achieved in a single fixed-bed reactor and steam addition can be significantly reduced. In a similar application RTI developed for a chemical conversion of syngas to Substitute Natural Gas (SNG), the transport-reactor based process resulted in a capital cost reduction of > 45% over the traditional multiple stage fixed-bed SNG production system (DE-FC26-06NT42760).

However, at the start of this project, development of this ATWGS process was still only conceptual. The strong alignment of this ATWGS process with the objective of reducing the overall production cost of H₂-rich syngas in this project resulted in a small task dedicated to evaluating the technical feasibility of this conceptual process. The strategy for successfully completing this technical feasibility assessment used experimental testing to establish target conversions achievable with existing commercial WGS catalysts and identify the research and develop (R&D) needs for an attrition-resistant and active fluid-bed WGS catalyst. In parallel, techno-economic analyses were used to evaluate technical and economic feasibility.

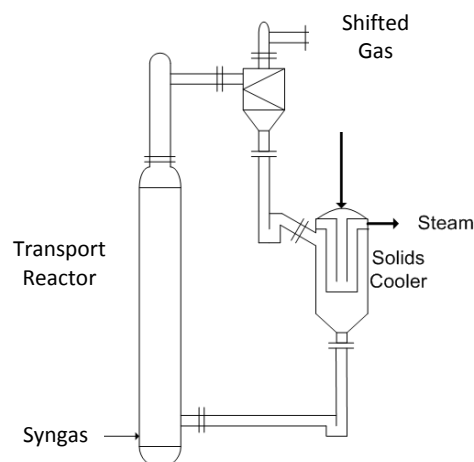


Figure 9. Conceptual schematic of transport reactor-based WGS process

4. IGCC Comparison Study

In this comparison study, a baseline or reference IGCC, which was based entirely on current commercial technologies and included 90% carbon capture, was selected from existing techno-economic analysis from DOE/NETL's "Advanced Systems Pathway" studies. The performance and cost structure of this reference plant were then compared with similar plants that incorporated the advanced technology processes. To allow the comparison of both the individual and potential synergistic benefits of these advanced technologies, three versions of the advanced plants were developed, which included a plant with each advanced technology separately and one with all the advanced technologies combined. The key technology components of the reference plant and each advanced plant are shown in **Table 3**.

Table 3. Case study matrix for IGCC with CO₂ capture

	Case 2a ¹	Case 2b	Case 2c	Case 2d
Gasification technology				
Shell gasification with lock hopper-based feed system	✓			✓
AR compact gasifier with dry solids pump (DSP) feed system		✓	✓	
Gas cleanup ²				
Two-stage Selexol™ for CO ₂ and sulfur removal	✓		✓	
RTI WDP with AACRP		✓		✓
Water gas shift				
Sour shift	✓		✓	
AFWGS		✓		✓
GE 7F advanced gas turbine	✓	✓	✓	✓
CO ₂ drying and compression (to 2,215 psia)	✓	✓	✓	✓

¹ Reference case based on Nexant's benchmark simulation of the NETL Report 1399 Case S1B.

² Selexol™ removes H₂S and CO₂. Additional trace contaminant cleanup technologies included as defined by DOE/NETL baseline studies.

The reference IGCC selected was Case S1B from the DOE/NETL's "*Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coals to Electricity*," May 2011, DOE/NETL 2012/1399 and incorporated the following technology components:

- A Shell gasifier with dry feed,
- Sour water gas shift process for CO conversion
- Dual stage Selexol™ acid gas removal (AGR) for sulfur and CO₂ capture,
- General Electric Advanced 7F turbines,
- Heat Recovery Steam Generator (HRSG),
- Steam turbine, and
- CO₂ compression and drying.

The net power output for this power plant was 472 MW. The specific reasons Case S1B and in particular the Shell gasifier was chosen as the reference case included:

- The Shell gasifier-based IGCC utilizes a pressurized lock hopper dry feed system. This provides a more direct comparison against AR's DSP feed system, which is part of its overall advanced gasification package. Because the reference and AR advanced gasification systems are based on dry coal feed, a more direct comparison can be drawn between the two, with minimal secondary integration impacts associated with differences in ancillary equipment/subsystems.

- Of all the gasification technologies evaluated in the DOE/NETL Baseline Studies, the Shell gasification-based IGCC has the highest efficiency. If the IGCC utilizing AR/RTI technologies can perform better than the Shell IGCC design in terms of efficiency and cost, it will clearly establish the integrated AR advanced gasification and RTI advanced syngas cleanup system as the best available IGCC technology.
- Shell's gasification technology is considered mature, having already been proven on a commercial scale. Therefore, overall performance and cost can be estimated at a higher confidence level.

As a basis of this comparison study, Nexant developed an ASPEN Plus model of the reference case to create the process heat and mass balance (HMB), which enabled estimation of the overall utility balance including the heat recovery in the HSRG, power generation from the steam cycle, as well as cooling water load breakdown. These values were then used to benchmark Nexant's model against DOE/NETL's reference case and demonstrate that the results were within a reasonable range of accuracy.

With the HMB for the reference plant, AR and RTI had the necessary data for input and target output values with which to develop the HMBs for their individual advanced technology processes. The results from the HMBs for the advanced technology processes were then fed back into Nexant's plant model to evaluate the performance for each advanced plant configuration.

For AR's advanced gasification system, the different components of their process are fairly well integrated. Their primary input is the coal feed, which based on the reference case selection was fully compatible with AR's dry feed advanced gasification process. Once AR confirmed that the specific coal processing and drying requirements in the reference case would work with the equipment in their advanced gasification process, not additional integration efforts were required.

RTI's advanced warm syngas cleaning technology processes are integrated through the process between the generation of the syngas in the gasifier and consumption of the H₂-rich syngas in the combustion turbines. As a result, the integration of RTI's advanced warm syngas cleaning processes was a little more challenging. In an effort to simplify this integration effort, the process block for which RTI provided information was expanded to cover all syngas processing between the syngas scrubber and the humidification prior to the combustion turbine. With control of all these individual process pieces, RTI could effectively integrate the operation of the individual processes to achieve optimal conversion, separation, and thermal efficiencies for the total integrated process. With regards to the integration of the individual process pieces within RTI's overall process package, this approach saved a significant amount of time and yielded a better solution.

Although this approach worked well for the processes within RTI's overall process package, it did require several iterations to effectively optimize the amount of WGS conversion coupled with the AACRP to maximize the power generated in the combustion and steam turbines while meeting the 90% carbon capture requirement. The specific selectivity for removal of CO, CO₂, and H₂ in the amine solvent is different than in Selexol™. Consequently, the WGS conversion had to be adjusted to optimize the relative amounts of H₂ and CO to minimize the loss of these components in AACRP, while still achieving 90% carbon capture and optimal power generation.

Because RTI's process package was effectively sandwiched between the gasifier and turbine processes, it had both inlet and outlet specifications that the process had to achieve. The only specification that was not effectively optimized was pressure. Pressure drop across a process is very equipment specific, thus conceptual design processes use estimations for calculating the pressure drop across a process. However, as AACRP is based on a chemical solvent rather than a physical solvent, operating pressure will have an impact on this process. The specific impact of concern for this IGCC application was the auxiliary load required to drive the pumps for AACRP. Higher pressure drives this load up as a major fraction of the CO₂ was regenerated by pressure swing and only a small polishing thermal regeneration was used to minimize the thermal requirements. For cases in which the pressure drop was not optimized for AACRP, the auxiliary load for the pumps would be higher than necessary. In general, this impact would be small (< 5 MW), but with the small differences between some of the different advanced plants, this did have an impact.

The cost estimation methodology followed the DOE/NETL's Series of Quality Guidelines for Energy System Studies (QGESS). Cost estimation for the advanced plants effectively broke down into two groups. The first group was all the equipment that was common between the reference and advanced plants. This essentially was all the equipment that was not part of AR's or RTI's advanced technology processes. For this equipment, the cost was scaled based on the guidelines and scaling information provided in DOE/NETL's Capital Cost Scaling Methodology document.

For the advanced technology processes, the costs were developed specifically for these processes by either AR and Nexant or RTI. For AR's advanced gasification process, AR provide the cost of their specific proprietary equipment based on their cost associated with construction of their pilot plant facilities. For the more generic equipment in the gasifier process, Nexant developed a bottom-up cost estimate based on sizing and their commercial experience.

For RTI's advanced syngas cleaning technology processes, RTI developed a bottom-up cost based on equipment sizing and the actual cost from their 50 MWe pre-commercial demonstration unit installed at TEC's Polk Power Station.

The specific contingencies used for the different technology components of the advanced technologies used in this study are shown in **Table 4**. The development of these contingency values was based on the criteria recommended by the Association for Advancement of Cost Engineering International (AACE) as specified in DOE/NETL's *Cost Estimation Methodology for NETL Assessment of Power Plant Performance*.

Table 4. Process contingencies for advanced technologies being evaluated

Account number	Advanced technology	Process contingency (% of associated process capital)	Rationale
2.3	AR DSP	20%	The AR DSP and static splitter feed system are currently demonstrated at the small pilot level. The system shares similar types of commercially demonstrated equipment with the conventional lock hopper feed system e.g., feed bins, compressors.
4.1	AR compact gasifier	25%	At 25%, this is within range of contingencies applied to technologies demonstrated at the small pilot level
5A.1	AACRP	20%	This is consistent with the contingency applied for similar processes such as Selexol™ and Rectisol® in the DOE/NETL studies.
5A.2	RTI DSRP	20%	The DSRP technology has been demonstrated in RTI's pilot testing at Eastman Chemical Company (DE-AC26-99FT40675)
5A.4a	RTI WDP	20%	The WDP technology has been demonstrated in RTI's 50 MWe demonstration plant project (DE-FE0000489)
5A.4b	RTI AFWGS	0%	Fixed-bed sweet water gas shift technology is commercially proven

The results from this cost estimation methodology for the reference IGCC case and the three advanced IGCC cases are shown in **Table 5**. More details about the process, performance and cost breakdown for each case is provided in the comprehensive techno-economic analysis (TEA) report provided as **Appendix A**.

Table 5. IGCC results summary

Case	Case 2a	Case 2b	Case 2c	Case 2d
IGCC configuration				
Gasifier	Shell	AR	AR	Shell
Sulfur and CO ₂ removal	Selexol™	RTI WDP	Selexol™	RTI WDP
CAPEX, \$MM				
Total Installed cost (TIC)	\$1,521	\$1,242	\$1,309	\$1,444
Total plant cost (TPC)	\$2,030	\$1,674	\$1,772	\$1,919
Total overnight cost (TOC)	\$2,487	\$2,048	\$2,174	\$2,344
OPEX, \$MM/yr (100% capacity factor basis)				
Fixed operating cost (OC _{fix})	\$74.2	\$62.8	\$65.9	\$70.6
Variable operating cost, less fuel (OC _{var})	\$56.9	\$49.1	\$50.5	\$55.2
Fuel (OC _{fuel})	\$50.4	\$48.4	\$48.4	\$50.4
Total OPEX	\$181.5	\$160.2	\$164.8	\$176.2
Power production, MWe				
Gas turbine	430.0	429.9	430.2	429.6
Steam turbine	224.1	211.3	209.3	226.4
Auxiliary power consumption	193.4	178.7	177.8	192.2
Net power output	460.6	462.5	461.8	463.8
Emissions				
SO ₂ emissions (lb/MWh _{gross})	0.0063	0.0013	0.0053	0.0014
SO ₂ emissions (lb/MMBtu)	0.0008	0.0002	0.0007	0.0002
Fuel rate and efficiency				
Coal feed rate, tpd AR coal	7,032	6,752	6,752	7,032
Net efficiency, %	31.32%	32.75%	32.70%	31.53%
Power generated, MWh/yr (MWh)	4,034,901	4,051,626	4,045,045	4,062,618
TOC CAPEX, \$/kW	5,400	4,428	4,709	5,054
Total OPEX, \$/MWh	44.97	39.55	40.74	43.36
COE, excl CO ₂ TS&M, mills/kWh	145.3	122.0	128.3	137.3
COE, incl CO ₂ TS&M, mills/kWh	166.2	142.0	148.4	158.1
Cost of CO ₂ avoided excl CO ₂ TS&M, \$/ton CO ₂	79.8	51.1	59.0	70.0
Cost of CO ₂ avoided incl CO ₂ TS&M, \$/ton CO ₂	105.1	75.2	83.1	95.0

Table 2 clearly quantifies the benefits associated with the integration of AR advanced gasification and RTI's advanced syngas cleanup technologies in an IGCC with > 90% carbon capture. Comparison Case 2b, in which these two advanced technologies were integrated achieved the best overall performance of all comparison cases completed in all the following categories:

- Overall plant efficiency: 32.75%
- Capital cost: \$2,048 MM
- Total operating cost: \$160.2 MM
- COE: 122.0 mills/kWh

- Cost of CO₂ avoided: 51.1 \$/ton CO₂
- SO₂ emissions: 0.0013 lb/MWh_{gross}

The success achieved through integration of these advanced technologies comes from the impressive benefits associated with each individual technology. The AR advanced gasification system, which is comprised of DSP and AR's compact gasifier, had a TPC of \$313 MM. Compared with the TPC for the Shell gasification system of \$510 MM, AR's advanced gasification system only costs 61% as much as Shell's gasification system. This difference represents about a 12 mills/kWh cost of electricity (COE) advantage for the AR gasification system compared with Shell's gasification system.

RTI's advanced syngas cleanup system also provides a significant reduction in capital costs when compared with a dual-stage Selexol™ system. From Case 2d, the TPC for RTI's advanced syngas cleanup system, which includes syngas desulfurization, sulfur recovery, AFWGS, CO₂ capture and compression, was \$252 MM. The TPC for the corresponding system based on Selexol™ was \$341 MM. This results in an \$89 MM capital cost advantage (26% capital cost reduction) and about a 6 mills/kWh advantage in COE.

Additional synergistic benefits are obtained in Case 2b when AR's and RTI's advanced technologies are integrated. In Case 2b, the TPC for the syngas desulfurization, sulfur recovery, WGS, CO₂ capture and compression is \$246 MM. The corresponding system in Case 2a (Reference case) has a TPC of \$341 MM which results in a \$95 MM capital cost advantage for WDP over the Selexol™ process. This represents an additional capital cost advantage for integration of both advanced technologies of \$6 MM.

From these conclusions, the integration of AR advanced gasification and RTI advanced syngas cleanup processes results in the following benefits when compared with the best available commercial technologies:

- ~ 1.4% net efficiency improvement,
- ~18% reduction in capital cost per kW of capacity,
- ~12% reduction in the operating cost per kWh generated, and
- ~ 75-79% reduction in sulfur emissions.

Because of the significant economic and efficiency benefits, market forces will drive commercial implementation of these technologies to realize the economic benefits. The fact that integration of these two technologies optimizes the benefits will also drive integrated deployment of these technologies. The fact that the RTI advanced syngas cleanup process also reduces sulfur emissions creates a condition where market forces will also drive the implementation of cleaner technologies.

Another component of this techno-economic analysis was sensitivity analysis. The sensitivity of the COE was evaluated as a factor of plant capacity factor, coal price, CO₂ sales price and CO₂ emissions costs. In all cases, the advanced plants had lower COE than the reference case. The sensitivity analysis also compared the capital for the entire IGCC plant as a function of the cost of each advanced technology. This sensitivity analysis showed that the benefits in IGCC capital cost for AR's gasification and RTI's WDP technologies still exists for a 30% increase in the total plant cost for these technologies. All of these factors indicate that when ready for commercial deployment, these technologies will be favored over existing commercial technologies.

5. Coal-to-Methanol Comparison Study

The selection of the reference case for the Coal-to-Methanol (CTM) comparison study was simple as there was only one case in DOE/NETL “*Baseline Analysis of Crude Methanol Production from Coal and Natural Gas* revised July 29, 2014, DOE/NETL, 341/020514” that incorporated 90% carbon capture. Fortunately, this case was based on Shell gasifier technology with dry feed, which allowed the same comparison benefits for the gasifier technology identified in the IGCC study. A unique aspect of this study was that a natural gas combined cycle (NGCC) with 90% carbon capture was included in the overall plant to ensure a power neutral plant with zero export/import of power and 90% carbon capture. The specific methanol production from this CTM plant was 10,000 metric tons per day. The specific commercial technologies included in this reference case were:

- A Shell gasifier with dry feed,
- Sour water gas shift process for partial CO conversion
- Dual stage Rectisol® AGR for sulfur and CO₂ capture,
- Natural gas turbine and Fluor Econoamine CO₂ capture,
- HRSG, and
- Steam turbine, and
- CO₂ drying and compression.

As for the IGCC comparison case, this CTM comparison case study was based on three advanced plant configurations. Two of these advanced plant configurations incorporated only one of the advanced technologies. The third advanced plant incorporated both advanced technologies permitting the identification of potential for synergistic benefits. The specific component includes in each case are provided in **Table 6**.

Table 6. Case study matrix for crude methanol production with CO₂ capture

	Case 3a ¹	Case 3b	Case 3c	Case 3d
Gasification technology				
Shell gasification with lock hopper-based feed system	✓			✓
AR compact gasifier with dry solids pump (DSP) feed system		✓	✓	
Gas cleanup				
Rectisol® for CO ₂ and sulfur removal ²	✓		✓	
RTI WDP with AACRP		✓		✓
Water gas shift				
Sour shift	✓		✓	
AFWGS		✓		✓
Methanol production	✓	✓	✓	✓
NGCC power generation with fluor econoamine CO ₂ capture	✓	✓	✓	✓
CO ₂ drying and compression (to 2,215 psia)	✓	✓	✓	✓

¹ Reference case based on Nexant’s benchmark simulation of the Crude Methanol Study Case 2.

² Rectisol® removes H₂S and CO₂. Additional trace contaminant cleanup technologies included as defined by DOE/NETL baseline studies.

The same approach used for the IGCC comparison study was used for this CTM comparison study. However, there were two main departures from the design constraints established by the reference case. During Nexant’s analysis of the reference case, they found that the two Rolls-Royce Trent 60 natural gas-fired turbines did not generate an outlet temperature with sufficient latent heat to

superheat/reheat all the steam raised in the gasification and methanol synthesis processes for power generation in the steam turbine. Nexant's solution was to use three General Electric MS6001B gas turbine each producing 45 MWe, which is similar to the power generation approach used in DOE/NETL's "*Cost and Performance Baseline for Fossil Energy Plants Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis*, DOE/NETL—2011/147".

The second difference was that the advanced plant configurations resulted in the generation of additional power in excess of the auxiliary load. This power was exported. Thus, the advanced plant configuration did ensure that all the CTM plant's auxiliary load was supplied for energy for which a 90% carbon capture criteria was met.

The results from this cost estimation methodology for the reference CTM case and the three advanced CTM cases are shown in **Table 7**. More details about the process, performance and cost breakdown for each case is provided in the comprehensive TEA report provided as **Appendix B**.

Table 7. CTM results summary

Case	Case 3a	Case 3b	Case 3c	Case 3d
CTM configuration				
Gasifier	Shell	AR	AR	Shell
Sulfur and CO ₂ removal	Rectisol®	RTI WDP	Rectisol®	RTI WDP
CAPEX, \$MM				
Total installed cost (TIC)	\$3,581	\$2,732	\$2,897	\$3,386
Total plant cost (TPC)	\$4,802	\$3,707	\$3,950	\$4,513
Total overnight cost (TOC)	\$5,892	\$4,567	\$4,857	\$5,548
OPEX, \$MM/yr (100% CF basis)				
Fixed operating cost (OC _{fix})	\$162.9	\$128.0	\$135.6	\$153.7
Variable operating cost, less fuel (OC _{var})	\$103.9	\$95.3	\$87.9	\$111.2
Coal feedstock (OC _{Coal})	\$259.2	\$244.8	\$244.8	\$259.2
Natural gas feedstock (OC _{NG})	\$85.4	\$77.6	\$78.4	\$85.4
Power import/(export)	\$0.4	(\$14.1)	(\$4.0)	(\$12.9)
Total OPEX	\$611.8	\$531.4	\$542.7	\$596.6
Methanol production, mtpd (100% CF basis)				
Gross methanol production	10,218	10,079	10,204	10,100
Rectisol® methanol makeup	8	0	6	0
Net methanol production	10,210	10,079	10,198	10,100
NGCC plant power generation, MWe				
Gas turbine	127.1	127.9	127.8	127.1
Steam turbine	264.7	239.2	199.1	292.8
Total power generation	391.8	367.1	326.8	419.9
Auxiliary consumption, MWe				
Process plant				
Coal handling and drying	17.4	16.4	16.4	17.7
Gasification and ASU	185.3	163.6	165.1	184.3
Process AGR	42.5	42.6	31.7	39.1
Process CO ₂ compression	65.2	64.8	52.4	75.6
Methanol synthesis	27.3	4.7	5.0	23.2
Process plant BOP	22.5	20.5	20.2	23.0

Benefits of Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production

Case	Case 3a	Case 3b	Case 3c	Case 3d
Total process plant auxiliary consumption NGCC plant	360.2	312.5	290.9	362.9
NGCC auxiliary consumption	22.4	23.5	17.7	28.4
NGCC PCC	3.9	4.1	4.1	3.9
NGCC PCC CO ₂ compression	6.0	w/process	6.4	w/process
Total NGCC AUXILIARY CONSUMPTION	32.3	27.6	28.2	32.2
Total auxiliary consumption	392.6	340.1	319.1	395.1
Net power generation/(consumption)	(0.8)	26.9	7.7	24.8
Fuel rate and efficiency				
Coal feed rate, tpd AR Coal	19,418	18,339	18,339	19,418
Natural gas feed rate, MMBtu/day	38,152	34,675	35,051	38,152
Total thermal input In (LHV), MMBtu/day	354,934	333,984	334,323	354,934
Methanol fuel thermal output (LHV), MMBtu/day	188,547	185,730	188,421	185,877
Power output, MMBtu/day	(63)	2,207	634	2,029
Thermal efficiency (LHV), %	53.1%	56.3%	56.5%	52.9%
CO ₂ capture				
CO ₂ capture product, tpd	19,413	18,201	17,576	20,021
CO ₂ vented, tpd	2,966	2,359	2,833	2,538
Carbon emitted as CO ₂ , % carbon in fuel	7.9%	6.6%	7.9%	6.7%
Methanol RSP				
Methanol RSP, excl CO ₂ TS&M, \$/ton ²⁵	424.2 500.7	347.3 407.3	359.5 422.7	408.9 481.7
Incl CO ₂ TS&M, \$/ton ²⁵	441.5 517.9	363.6 423.7	375.1 438.3	426.8 499.6
RSP components details, \$/gal				
Capital ²	0.90 1.15	0.70 0.91	0.74 0.95	0.85 1.10
Fixed O&M	0.15	0.12	0.12	0.14
Variable O&M	0.08	0.08	0.07	0.09
Coal	0.21	0.20	0.20	0.21
Natural gas	0.07	0.06	0.06	0.07
Power	0.00	(0.01)	(0.00)	(0.01)
CO ₂ TS&M	0.06	0.05	0.05	0.06
Total ²	1.46 1.72	1.21 1.40	1.24 1.45	1.41 1.66
Cost of CO ₂ captured, \$/tonne CO ₂ ²³	18.8 20.9	18.9 21.1	18.7 21.0	19.0 21.1
Cost of CO ₂ avoided, \$/tonne CO ₂ ²⁴	29.8 31.9	29.9 32.1	29.7 32.0	30.0 32.1

¹ Capacity factor assumed to be 90%.

² Values shown are for two financial structures. The first (lower value) is based on the loan guarantee finance structure. The second (higher value) is based on the commercial fuels finance structure.

³ Excludes CO₂ TS&M.

⁴ Includes CO₂ TS&M.

⁵ Based on 301.73 gallons/ton.

Figure 10 illustrates the breakdown of the cost of production. It can be seen from Figure 10 that the cases differ most significantly in terms of capital cost on a \$/gal basis.

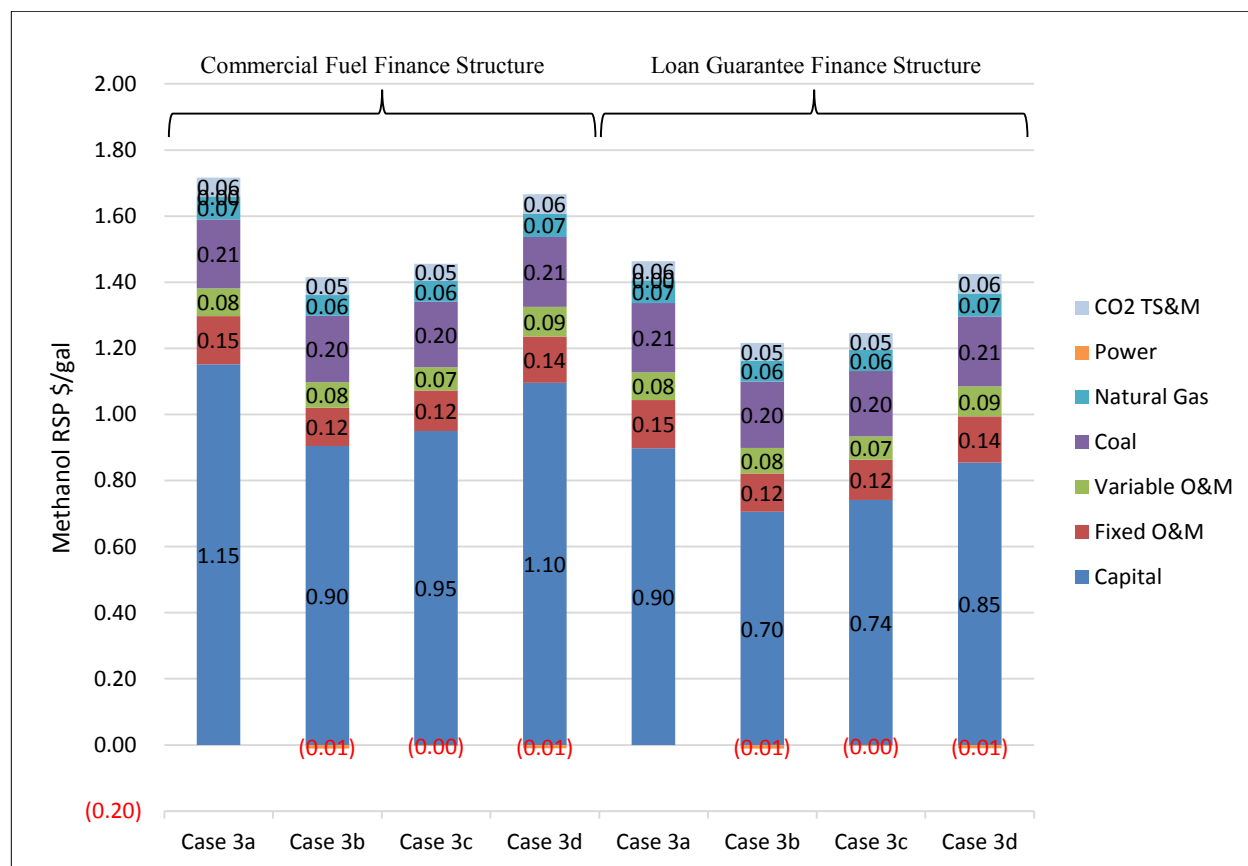


Figure 10. Components of methanol RSP

This breakdown by component of the methanol RSP for the different cases illustrates the relative contribution of each component. In all cases, the capital cost accounts for 60% or more of the methanol Required Selling Price (RSP). Both the AR gasification and RTI advanced warm syngas cleanup technologies significantly reduce capital cost and methanol RSP. The AR gasification system reduces the capital cost of the overall gasification block (including the dry feed system) by 40–50% and results in about \$855 MM or an 18% decrease in TPC of the entire CTM plant (including benefits to other blocks due to efficiency improvements). The RTI advanced warm syngas cleanup system reduces capital costs of the overall gas cleanup block (including AGR, SRU, and WGS,) by 30–40% and results in reduction in TPC of the entire CTM plant by about \$280 MM or about 6%. These capital cost reductions result in a 16% and 4% decrease in methanol RSP for the AR gasification and RTI advanced warm syngas cleanup technologies, respectively.

The inherent ability of the AR DSP and gasification systems to operate at higher pressures leads to a number of synergistic advantages that result in additional capital cost reductions. The benefits include:

- Elimination of a syngas compressor to boost the syngas pressure for the methanol synthesis process,
- Reduction of the capital cost for the two-stage Rectisol® AGR, and
- General reduction in equipment size throughout the syngas generation and cleanup processes associated with the reduced volumetric flow at higher pressures.

The unique aspect of RTI's advanced warm syngas cleanup is the decoupling of the sulfur and CO₂ removal processes. The cost reduction of this decoupling was discussed in the previous section and resulted in a 30–40% capital cost reduction in the overall cleanup block or \$280 MM or 6% of the TPC of the entire CTM plant. However, RTI's advanced warm syngas cleanup technology also results in

reductions in overall emissions. The sulfur emissions for Case 3a (reference case) and 3c (AR gasifier and Rectisol® AGR) are 0.035 and 0.040 lb of SO₂/MMBtu of coal, respectively. For both cases based on RTI's advanced warm syngas cleanup (Cases 3b and 3d), the sulfur emissions are < 0.0001 lb of SO₂/MMBtu of coal. This represents over a 99% reduction in sulfur emissions vs Reference Case 3a. The cases with RTI advanced warm syngas cleanup also reduce net CO₂ emissions by between 380 tpd and 420 tpd, which is equivalent to about a 13–15% reduction in CO₂ emissions. The net result of this decoupling of sulfur and CO₂ removal is a substantial reduction in the cost of contaminant removal allowing the reduction of emissions to be near zero.

The Case 3a reference plant consumes about 0.8 MW of net power. Because of auxiliary load reductions and/or improved steam generation from process heat recovery, the case studies with AR and/or RTI advanced technologies result in net power generation. For both cases with RTI's advanced warm syngas cleanup, sufficient steam is recovered from the process to generate ≥ 24 MW of net power. This illustrates the benefits of the higher overall process efficiencies enabled by RTI's WDP process. Case 3c with AR's gasification system generates about 8 MW of net power.

The various CTM plant cases show that at the lower operating pressure used in Cases 3a (reference case) and 3d (Shell gasifier with RTI's advanced warm syngas cleanup), auxiliary loads associated with the chemical solvent used in AACRP are less than those associated with the physical solvent used in Rectisol®. At the higher operating pressure enabled by AR DSP technology, the auxiliary load for the physical solvent-based process (Rectisol®) becomes less than for the chemical solvent based process (AACRP). Although the difference is less than 6% of the total auxiliary load for the CTM plant, this difference illustrates the importance of considering operating pressure in the criteria for CO₂ capture technology selection. In this case, it might have been interesting to see how a coupling of RTI WDP with Rectisol® would have differed from WDP with AACRP.

Because of the increased amount of CO₂ capture for the cases with RTI advanced warm syngas cleanup, the auxiliary loads associated with the CO₂ compression are higher. The overall reductions in auxiliary loads and increased power generated from process heat recovery for the cases with RTI advanced warm syngas cleanup result in over 24 MW_e of net power generation, which more than offsets the increased auxiliary load associated with the additional CO₂ compression.

The magnitudes of the cost advantages found in this study reflect the specific constraints associated with this study. The reference case selected for this study was based on Case 2 in the Crude Methanol Study. One of the key design constraints in Case 2 was the decision for the overall CTM plant to be essentially power neutral and for any power generated to also include > 90% carbon capture on its emissions. This decision, while allowing for a clear demonstration of > 90% carbon capture for the methanol product, forces the inclusion of a dedicated NGCC power plant with carbon capture. This has several very negative impacts. If any excess power is generated, this excess power would probably be sold to the grid at a discount because any negotiated power purchase agreement would exploit the fact that the power production is an inherent by-product of the methanol process. The breakdown of the methanol RSPs and sensitivity analysis demonstrate that any excess power generation has essentially no financial value, whereas a decision in the base case to allow power needs to be purchased across the fence from the grid would have enabled the net power generated by the AR and RTI advanced technologies to offset power purchases and reduce the methanol RSP. The requirement to include power generation processes with associated carbon capture also increases the overall capital cost of the CTM plant. As the capital cost represents over 60% of the methanol RSP, the cost of the power generation system and supporting subsystems adversely inflate the capital cost of the CTM plant and methanol RSP and somewhat diminish the relative net impact of utilizing the AR and RTI advanced technologies. Unfortunately sectioning out the capital cost of the power generating equipment and the associated increase in capital cost of the supporting systems is not possible in the current study, because of integration between the process heat recovery and power generation system. Finally, the inclusion of power generation processes also increases the overall auxiliary load by about 8%.

In Case 2 from the Crude Methanol Study, the CTM plant has a total of 9 Shell gasifiers with eight in continuous operation and one spare. Based on the coal throughput, the size of these Shell gasifiers

is about 2,000 tpd. Based on the assumption that the reference case should reflect actual commercially operating plants, this gasifier selection is practical. However, this study focuses on evaluating emerging technologies which are expected to be demonstrated commercially in the next 5 years. Because of this approach, the AR gasification system included a 3,600 tpd coal gasifier. Shell's webpage also announces that they have designed and will supply a Shell gasifier capable of 3,200 tpd of coal. If Shell's larger gasifier had been used in the reference case, a total of 5 continuously operating gasifiers would have been required, which is identical to the number of AR gasifiers used in this study. There is no doubt that use of the AR gasification system would still result in a substantial reduction in capital cost (because AR gasifiers are estimated to cost substantially less than equivalent-sized Shell gasifiers), but the total magnitude of any cost reduction would be somewhat smaller.

As part of this study, Nexant estimated the performance and costs of the Rectisol® process operating at 815 psia versus 515 psia. It was determined that the Rectisol® process' cost as reported in the DOE/NETL Crude Methanol Study were within the lower end of Nexant's range of cost estimates. Therefore, the cost savings from utilizing RTI's advanced syngas cleanup system reported in this study are in all likelihood conservative.

The importance of operating pressure on the cost and performance of the CO₂ capture system was identified. Because RTI's advanced syngas cleanup system is integrated with the upstream gasification processes and the downstream methanol synthesis processes, there are many constraints that must be met. Although all of the heat and mass balance constraints were met, optimizing the pressure drop across this system was not a priority. As a consequence, additional optimization of the pressure drop across the RTI advanced syngas cleanup system could further increase economic performance benefits for this system.

Since one of the key benefits of RTI's WDP technology is that it decouples the sulfur and CO₂ removal processes, the selection of the CO₂ capture technology should be made based on economic and performance criteria. For integration of RTI's advanced syngas cleanup system and CO₂ capture in this study, an AACRP system was chosen based on the cost data for an integrated system available from RTI's 50 MW_e demonstration plant project (DE-FE0000489). Although this did take advantage of the best available cost information, it did not necessarily maximize the benefits of RTI's advanced syngas cleanup technology. In a future study, the potential benefits of integration of RTI's advanced syngas cleanup technology with Rectisol® for carbon capture would be an interesting case study, particularly for chemical production applications that would likely benefit from higher syngas pressures.

Several key findings were revealed by the sensitivity studies. The first finding was that AR's advanced gasification and RTI's advanced warm syngas cleanup technologies still had a lower methanol RSP than the reference case even when the equipment cost for these technologies was increased by 30%. The second finding was that the selling price for any excess electricity generated and sold had only a small (< \$10/ton) impact on the methanol RSP for a COE between 0 and \$120/MWh. The final observation was that individually AR's advanced gasification and RTI's advanced warm syngas cleanup technologies resulted in benefits over the reference case for every sensitivity study completed. However, the case with the most benefits in every single sensitivity study was the integrated case integrating both advanced technologies.

The substantial benefits from integration of AR's gasification and RTI's advanced syngas cleanup into a CTM plant, which include significant capital cost reductions, lower sulfur and CO₂ emissions, and improved energy efficiency, will generate market forces that will drive the implementation of the AR and RTI technologies to realize these benefits. All these factors indicate that when ready for commercial deployment, these technologies will be favored over existing commercial technologies.

6. Technical Feasibility Assessment of ATWGS

One of the key technical challenges for the ATWGS process is the development of fluidizable WGS catalysts. The key criteria for these fluidizable catalysts would be catalytic WGS activity, stability of catalyst activity, and attrition resistance. The experimental work completed under this project was limited to primarily testing WGS catalysts to establish their catalytic activity, stability, and attrition resistance. The first set of WGS catalysts tested included commercial WGS catalysts. Testing commercial WGS catalysts served to establish reasonable conversion levels for the TEA and benchmark activity levels for the fluidized-bed catalyst formulations. **Table 8** provides a summary about the three distinct types of WGS catalysts used commercially.

Table 8. Commercial WGS catalyst formulations

Catalyst	Trade name	Composition [wt%]	Application description
LTS (Low-temperature-shift)	ShiftMax 240®	59.0% CuO 30.5% ZnO 10.5% Al ₂ O ₃	This WGS catalyst is primarily used for final CO conversion to maximize the H ₂ concentration in the product stream. The typical operating temperature for the catalyst is 200°C to 250°C. This catalyst is extremely sensitive to sulfur poisoning.
HTS (High-temperature-shift)	ShiftMax 120®	89.5% Fe ₂ O ₃ 8.5% Cr ₂ O ₃ 2.0% CuO	This WGS catalyst is used for bulk conversion of the CO in syngas mixtures. This catalyst is more stable for high temperature operation which allows the higher conversion. It can tolerate up to 50 ppmv of sulfur, but is typically used with sulfur-free syngas
SGS (Sour-gas-shift)	ShiftMax 820®	10.0% MoO ₃ 4.0% CoO 18.0% MgO 68.0% Al ₂ O ₃	This WGS catalyst is typically used for bulk CO conversion in coal-derived syngas, because it requires at least 100 ppmv of sulfur in the syngas to maintain its catalytic WGS activity.

The second set of WGS catalysts tested included fluidized-bed formulations. To minimize catalyst development efforts on this project, catalyst preparation was limited to a maximum of two formulations for each type of commercial WGS catalyst. However, to maximize the benefits, selection of these materials was carefully made to permit evaluation of key preparation techniques or methods that would lead to increasing attrition resistance of the catalyst formulations. Subsequent R&D efforts outside of this project could be used for catalyst development leveraging the beneficial techniques identified in this project.

6.1 Description of Reactor Testing Systems

The performance of the WGS catalysts was determined in laboratory-scale microreactor systems. A schematic for the WGS test system is shown in **Figure 11**. A photo of the WGS microreactor systems are shown in **Figure 12**. The downflow fixed-bed reactor in these test systems consisted of a 0.37 in. ID stainless steel tube that contained a nominal 2 cm³. The reactor was heated with band heaters (Watlow) controlled by a thermocouple placed inside a thermowell in the catalyst bed. Mass flow controllers (Brooks 5850i) regulated delivery of the reaction gases from compressed gas cylinders to the system. Argon (Ar) was included in the gas stream fed to the reactor as an internal standard used in product analysis. The WGS catalysts were pretreated per manufacturer specifications before testing. After pretreatment, carbon monoxide (CO), CO₂, methane (CH₄), and hydrogen (H₂) were slowly added to reach the baseline syngas composition.

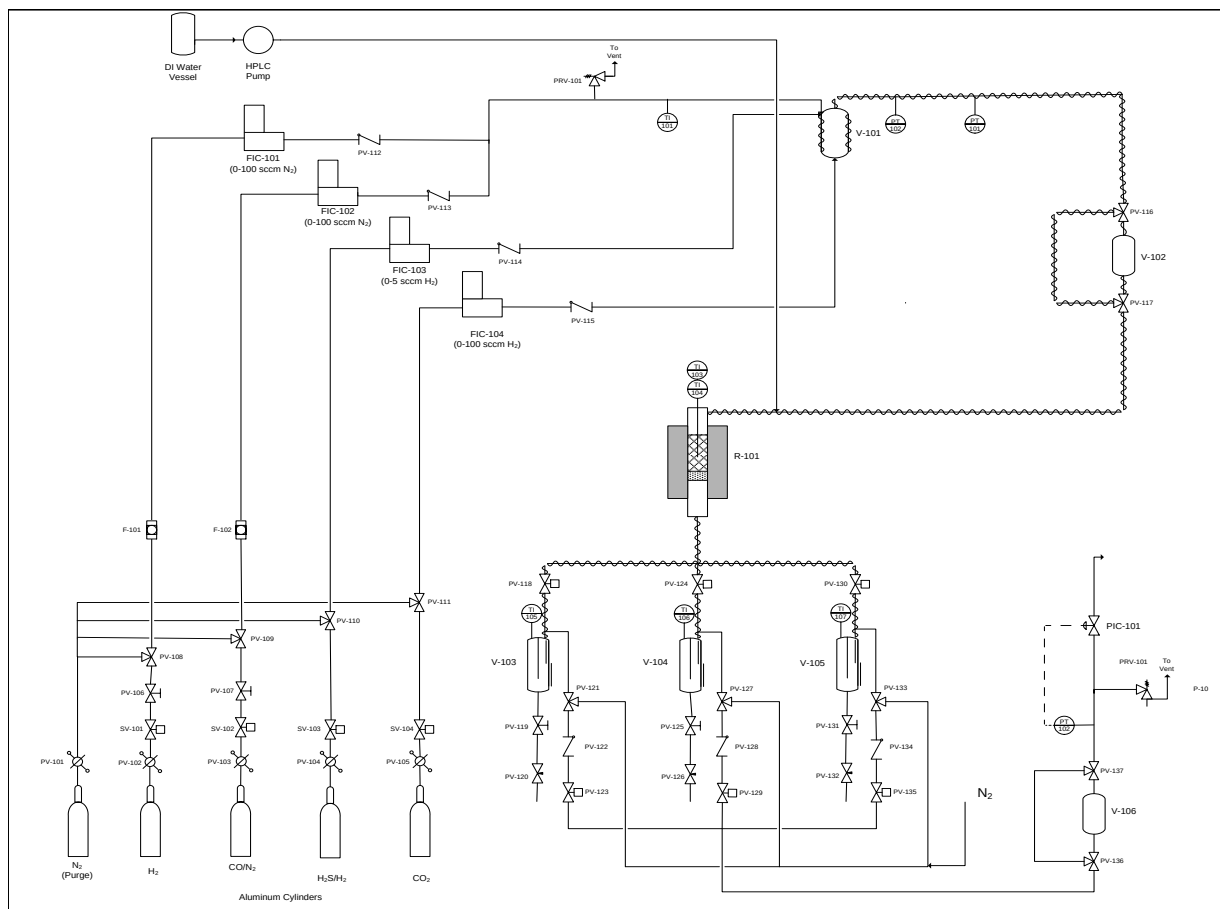


Figure 11. Schematic of WGS test system



Figure 12. Laboratory-scale microreactor systems for WGS testing

Product gases were analyzed by a Micro GC refinery gas analyzer (Agilent 3000). A cold trap at 10°C to 15°C collected any condensed water. Calibrations of analytical equipment were performed using calibration mixtures from Scott Specialty Gases.

6.2 Experimental Conditions

All the commercial catalysts came as fixed-bed pellets from the manufacturer. To prepare for testing, the pelletized catalysts were first crushed using a mortar and pestle and then sieved to collect particles in the size range of 53 to 104 μm . This particle size range was chosen to minimize gas channeling in the interparticle spaces within the catalyst bed and make them size compatible with the fluidized-bed WGS catalyst formulations to be tested. Because the WGS reaction is exothermic, α -alumina particles were added as inert material to reduce the reaction per volume in the catalyst bed and to increase the thermal mass available to adsorb this energy. The actions would help to promote more isothermal testing conditions in the catalyst bed. The particle size of the α -alumina used in the trials was about 125 μm to match the size of the catalyst particles.

Specific temperature, pressure, and baseline syngas compositions were selected for each WGS catalyst that are representative of commercial operations for each catalyst and are presented in **Table 9**. The conversion of CO resulting from the WGS reaction, $\text{H}_2 + \text{CO} = \text{H}_2\text{O} + \text{CO}_2$, was used to monitor WGS catalyst performance. A potential competing reaction to the WGS reaction is the methanation reaction, $3\text{H}_2 + \text{CO} = \text{H}_2\text{O} + \text{CH}_4$. Thus, selectivity for methane was also used to monitor the performance of WGS catalysts.

Table 9. Baseline WGS and FT system test parameters

Component [vol%]	Catalyst system		
	LTS	HTS	SGS
H ₂	36.1	17.7	21.3
CO	4.6	23.0	27.7
CO ₂	29	10.6	12.8
CH ₄	2.8	2.8	3.4
H ₂ O	27.5	45.9	34.9
H ₂ S	0	0	100 ppmv
Temperature [°C]	200	375	250
Pressure [psig]	500	500	500
Space velocity at STP [h ⁻¹]	5,000	5,000	5,000

6.3 Catalyst Testing Results

As attrition losses will be one of the key operating expenses for ATWGS, preparing attrition-resistant catalyst formulations was critical for demonstration of technical feasibility. Based on this specification for the catalyst formulations, the selection process focused on making very attrition-resistant materials either based on existing materials or preparation techniques that guaranteed high attrition resistance. In the end, we prepared two low-temperature shift (LTS) catalysts and one sour gas shift (SGS) catalyst. For the high temperature shift (HTS), we identified an in-house fluid-bed material that was developed for a different application, but had shown promising WGS activity during previous testing. After preparation of these materials, the first test to be conducted on these catalyst formulations was attrition resistance. Because bulk density is a required value for the calculation for evaluation of the Davison Index (DI) attrition resistance value, it was also measured as part of this process. Although we made two LTS catalysts, only the material with the best attrition resistance was allowed to move on into microreactor testing. The results from the attrition resistance testing are shown in **Table 10**. Table 10 shows that for all three WGS catalyst types, we were able to prepare catalyst formulations that had a DI of less than the typical values for an equilibrium catalyst for FCCs, which represent the standard for

commercial attrition resistance in fluid-bed applications. These catalyst formulations demonstrate that suitable attrition-resistant catalyst formulations can readily be produced.

Table 10. Attrition resistance characterization of fluid-bed catalyst formulations

Sample	Bulk density	Davison index	
	(g/cm ³)	Fines < 20 μ m	Fines < 40 μ m
LTS-1	1.64	1.7	5
HTS-1	2.05	1.4	3
SGS-1	1.00	4	9
E-cat (FCC)	0.96	6	15

The next step was to test the reactivity of these catalyst formulations. Target reactivity levels were established using the three commercial catalysts. The commercial catalysts were tested at typical commercial operating conditions, where their performance would be optimal. Furthermore, these tests extended over at least 100 hours of operation to capture any decline in reactivity. The results for the HTS catalyst formulations are shown in **Figure 13**. Figure 13 shows that the average CO conversion for the commercial catalyst was 78% and was 72% for the fluid-bed catalyst formulation (HTS-1). Although the activity of both the commercial and fluid-bed catalyst formulations both decrease over the first 50 to 100 hours of operation, the activity eventually becomes stable.

In addition to testing the effluent gas for the typical gases involved in WGS reaction, like CO, CO₂, and H₂, the effluent gas was also tested for methane. Methane is generated via the methanation reaction, which typically competes with the WGS reaction. For optimal performance, a WGS catalyst would be expected to have a very low selectivity for methane. **Figure 14** shows the test results for HTS-1. HTS-1 has essentially no selectivity for methane formation and there is no indication of any significant change in methane selectivity over the entire test.

In a second series of activity tests, the CO conversion was measured as a function of temperature. This demonstrated the acceptable operating ranges for the catalyst. As with the previous test, the concentration of methane in the effluent gas was measured to evaluate methane selectivity. **Figure 15**, which shows the results for HTS-1, indicates that below 400°C reactivity is kinetically controlled by catalyst activity. Above 400°C, conversion is limited by thermodynamic limitation of the WGS reaction.

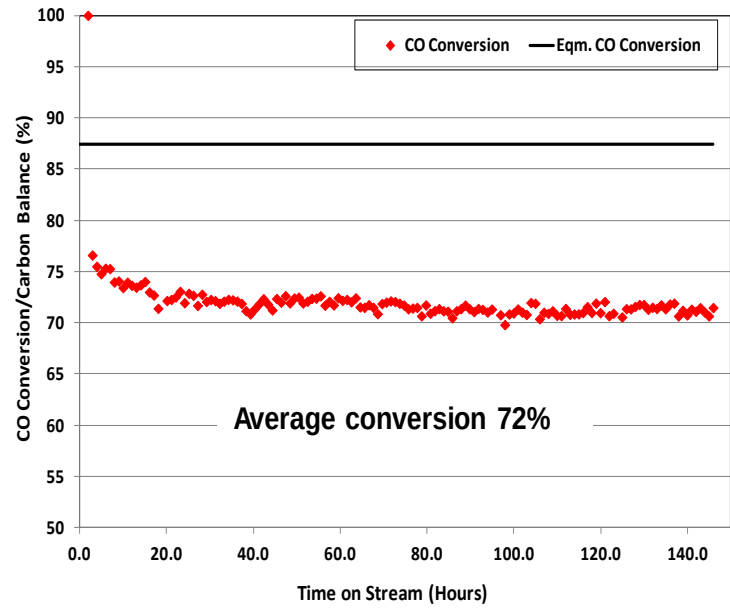
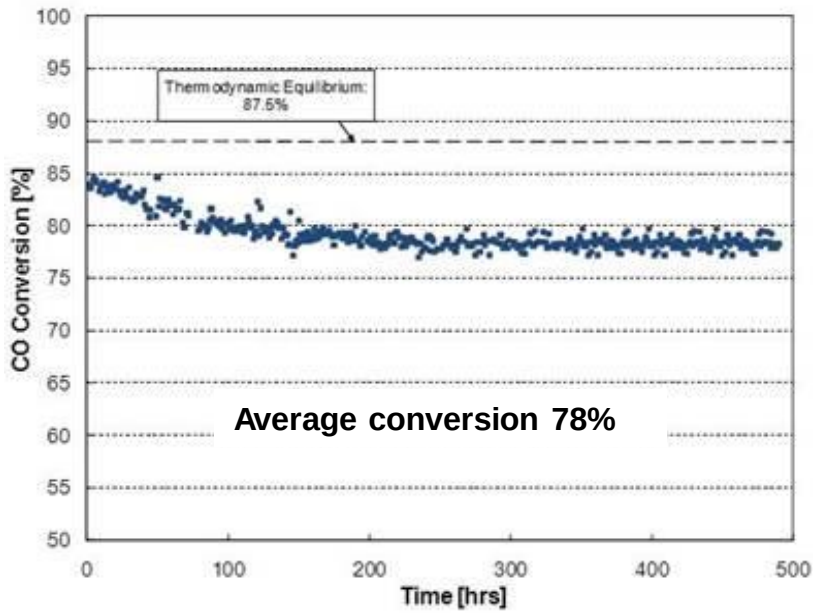


Figure 13. CO conversion performance from HTS catalyst formulations (commercial on left, RTI HTS-1 on right)

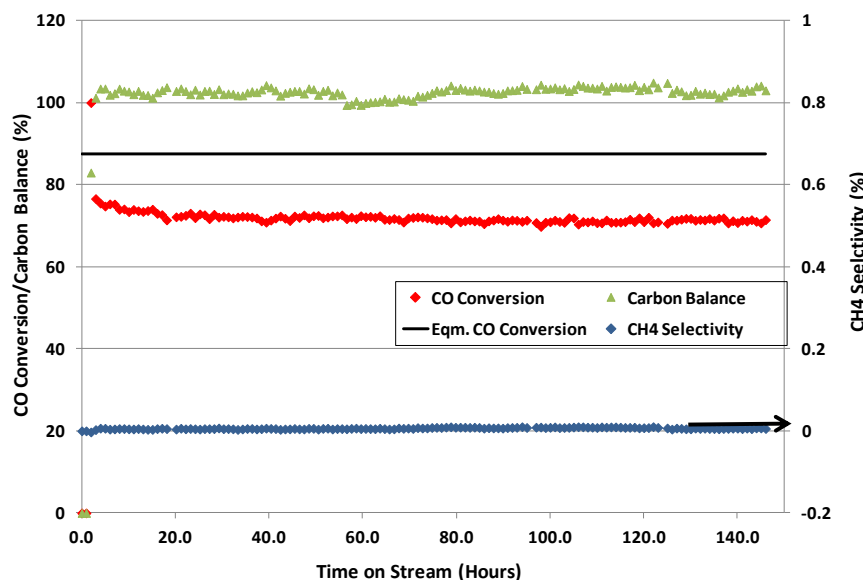


Figure 14. Methane selectivity of HTS-1

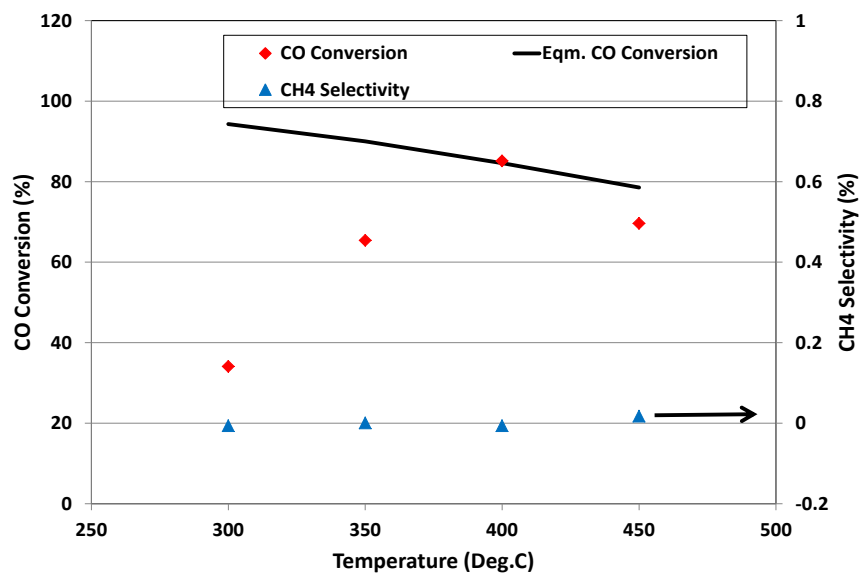


Figure 15. Effects of temperature on CO conversion of HTS-1

This same sequence of tests was also used for the SGS and LTS catalyst formulations. **Figure 16** shows the comparison of the CO conversion for the commercial SGS and fluid-bed SGS catalyst formulation (SGS-1). Unfortunately, SGS-1 was only able to achieve an average CO conversion of 35% compared to the commercial catalyst that had an average CO conversion of 78%. SGS-1 also showed significantly more deactivation over the 200 hours of testing than the commercial catalyst.

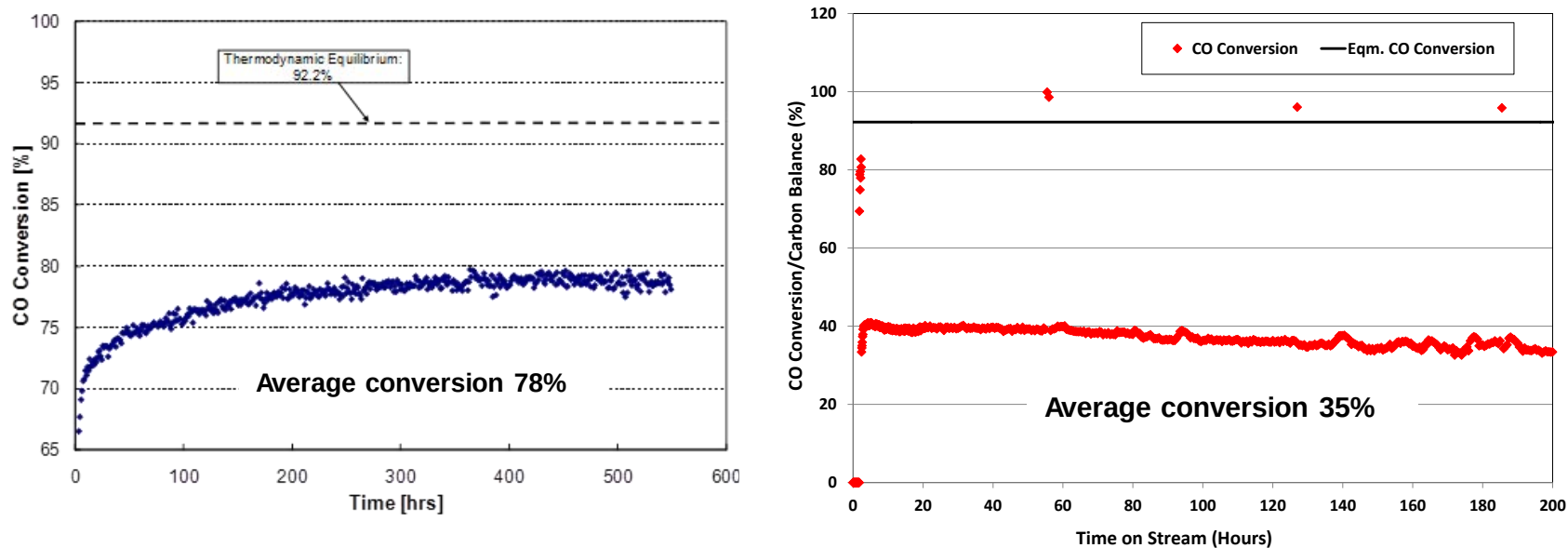


Figure 16. CO conversion performance from SGS WGS catalyst formulations (commercial on left, RTI SGS-1 on right)

The results from methane selectivity conducted as part of this test are shown in **Figure 17**. Although there is some methane formation, the methane selectivity is below 0.25% throughout the entire test, but did show a slow and steady increase.

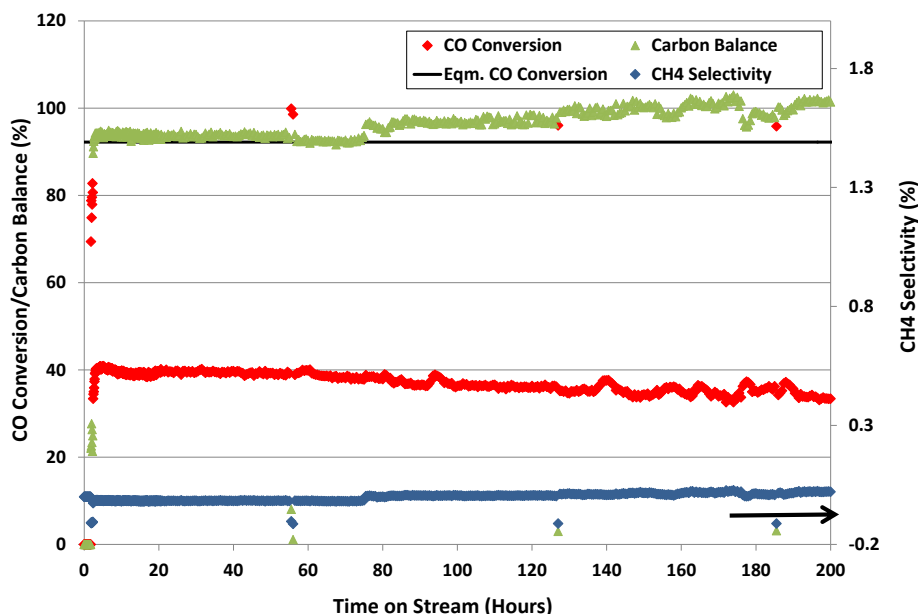


Figure 17. Methane selectivity for SGS-1

Parametric testing of the performance of SGS-1 at different temperatures, shown in **Figure 18**, demonstrates that below 350°C, the reaction is kinetically controlled. At 350°C, the reaction rate approaches the equilibrium CO conversion rate.

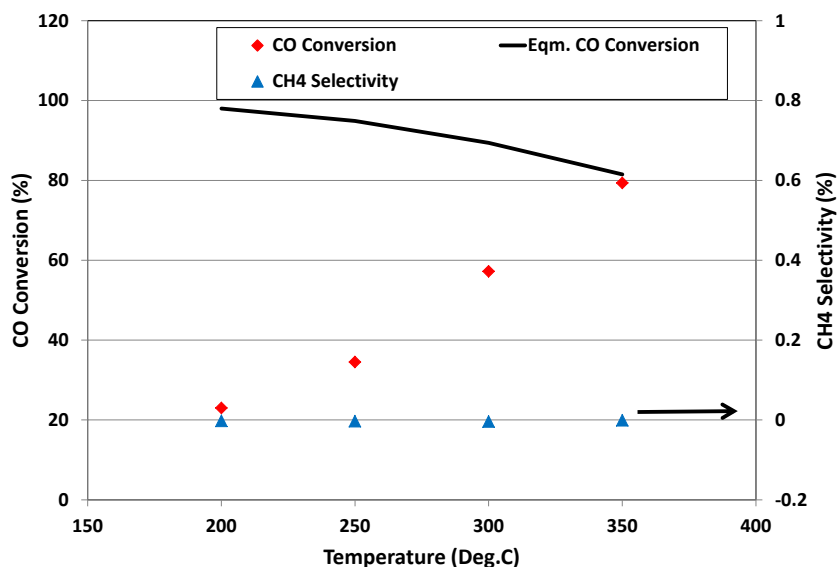


Figure 18. Effects of temperature on CO conversion of SGS-1

The activity comparison for the LTS catalysts are shown in **Figure 19**. The fluid-bed LTS catalyst formulation (LTS-1) had minimal LTS activity achieving an average CO conversion of only 20% in comparison with 87% for the commercial catalyst. Both the commercial and LTS-1 catalyst formulations show more decline in activity with time than the previous catalysts tested. Because LTS

catalysts must be extremely active because of their low operating temperatures, they also tend to experience more susceptibility to poisons and faster deactivation rates. Methane selectivity for LTS-1, shown in **Figure 20**, is still very low at $< 1\%$, but is higher than for either of the SGS-1 or HTS-1 catalyst formulations.

Parametric temperature testing of LTS-1 is shown in **Figure 21** and indicates that the reaction is kinetically controlled over the entire temperature range over which commercial LTS catalysts typically operate.

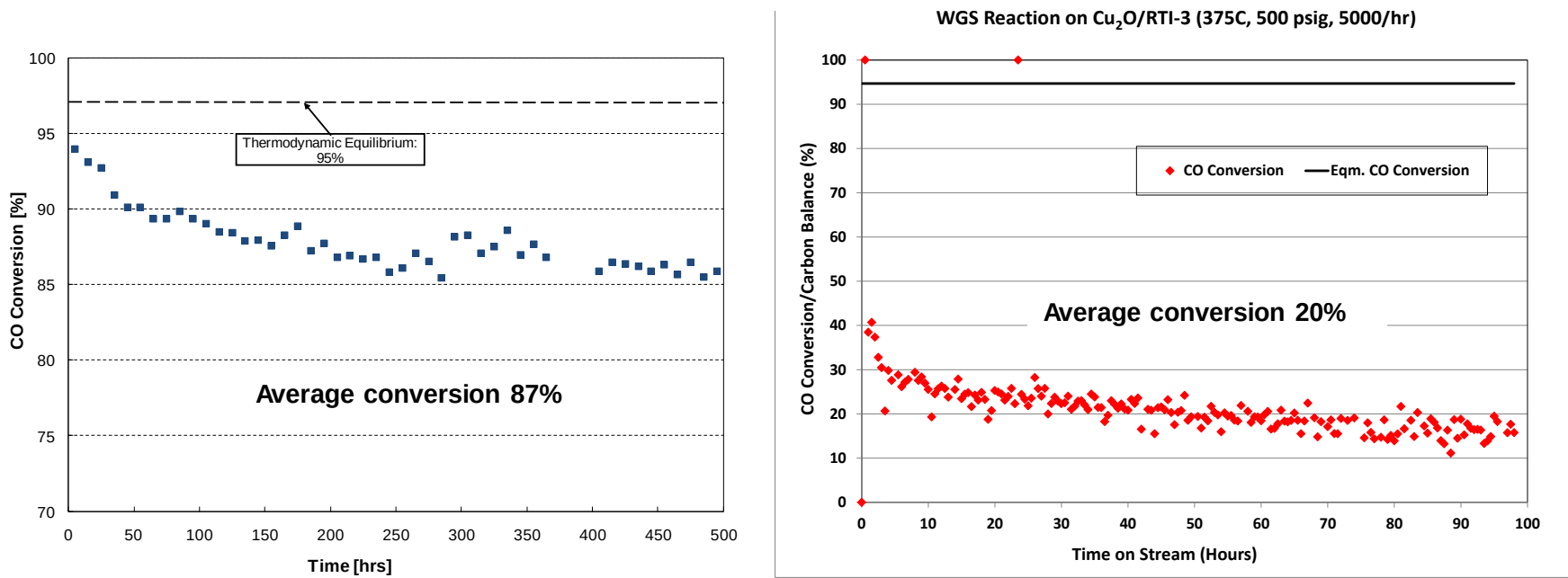


Figure 19. CO conversion performance from LTS catalyst formulations (commercial on left, RTI LTS-1 on right)

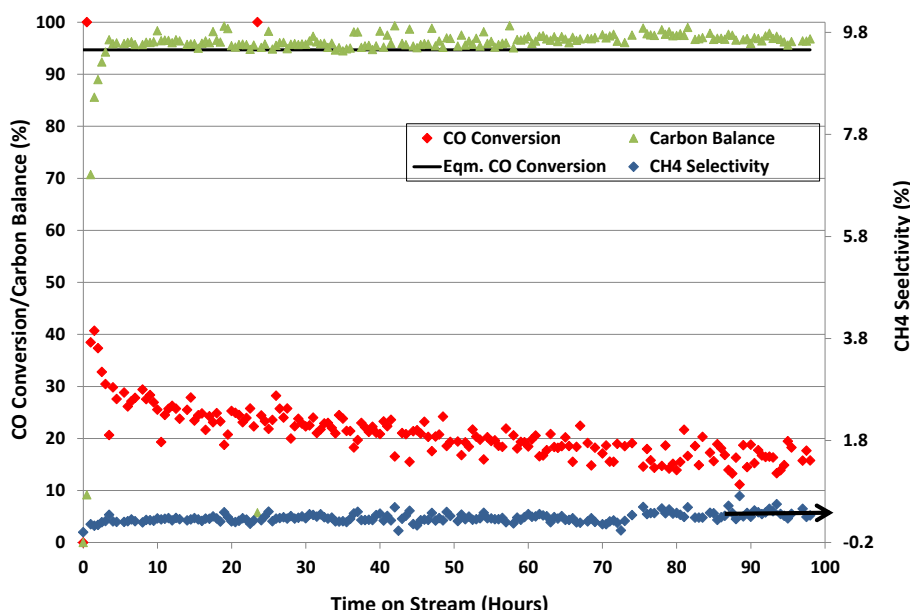


Figure 20. Methane selectivity for LTS-1

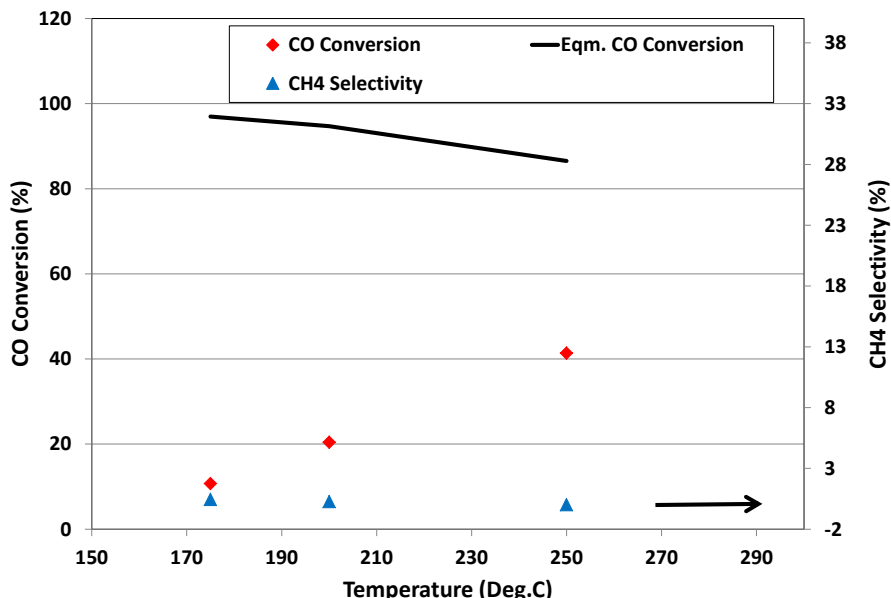


Figure 21. Temperature effects on CO conversion and methane selectivity for LTS-1

Although all the fluid-bed catalyst formulations prepared had exceptional attrition resistance, only the HTS-1 catalyst formulation also had a CO conversion that was comparable to the commercial catalyst. These characteristics coupled with its low methane selectivity make it an ideal starting catalyst formulation for ATWGS. SGS-1 represents a strong candidate for ATWGS for SGS applications. It had exceptional attrition resistance, a decent amount of CO conversion, and low methane selectivity. However, it would require some additional catalyst development to increase the WGS activity to match or exceed commercial SGS catalysts. As SGS applications represent a significant amount of the commercial WGS applications, this represents a very promising R&D area that should be considered for future funding.

6.4 Techno-Economic Analysis

Because DOE/NETL's pathway study case S1B from the DOE/NETL report "*Cost and Performance Baseline for Fossil Energy Plants Volume 3a: Low Rank Coal to Electricity: IGCC Cases DOE/NETL-2011/1399*," which incorporates a Shell gasifier using Powder River Basin (PRB) sub-bituminous coal, has served as the reference gasification technology and coal source for IGCC and methanol production studies completed in this project, the feed composition for the fixed-bed WGS system from the pathway study case was also selected as a basis for this techno-economic evaluation of ATWGS. However, rather than complete a full techno-economic analysis for an entire IGCC or CTM plant, the scope of this techno-economic analysis was narrowed to include just the WGS process.

With this narrower focus, the standard performance targets for the entire plant would not be valid. However, to complement and/or support the performance target of the entire plant, the specific performance targets chosen for this techno-economic analysis were the extent of CO conversion and amount of high pressure steam generated. Because the WGS system in pathway study S1B included two fixed-bed reactors, our base case also included two fixed-bed reactors. However, we did modify the heat recovery between the two reactors to maximize the amount of high pressure steam generated.

A standard bottoms up factored cost estimation methodology was implemented. This began with developing the heat and mass balances. As mentioned, the specific performance criteria used for these WGS processes were the extent of CO conversion and amount of high pressure steam generated. The heat and mass balances were then used to develop a sized equipment list and cost estimates for this equipment.

The breakdown for the CO conversion by reactor and HP steam generation is provided in **Table 11**. **Table 12** provides the costing information for each process.

Table 11. Performance results for water gas shift comparison cases

Criteria	Reference case	RTI ATWGS case
CO conversion (%)		
Reactor 1	80.7	91.1
Reactor 2	16.8	7.4
Total	97.5	98.5
HP Steam (lb/h)	254,059	269,907
BFW heated (lb/h)	341,327	341,327

Table 12. Cost estimates for water gas shift comparison cases

Equipment	Cost (2011\$)	
	Reference case	RTI ATWGS case
Reactor 1 (fixed)	2,217,200	
Reactor 1 (transport)		2,089,000
Reactor 2	904,200	468,900
Steam generator	4,580,000	1,840,000
BFW heater	423,400	706,700
Cross exchanger	2,560,000	341,327
Solids cooler		3,250,000
Total	10,984,400	8,335,540

One of the key goals of using ATWGS was to use circulation of the catalyst as a means to capture the exothermic heat of reaction. The performance results clearly demonstrate the value of this concept, with a significantly higher amount of CO conversion in the first reactor and the ability to generate more HP steam because of the heat extracted from the circulating catalyst particles. In addition to these performance improvements, ATWGS also resulted in about a 24% capital cost reduction.

6.5 Technical Feasibility Summary

The primary objective of this technical feasibility analysis task was to demonstrate that catalyst formulations with high attrition resistance and WGS activity could be readily prepared with scalable production techniques and potential economic benefits. The fluid-bed catalyst formulations prepared for each type of commercially practiced WGS process had an attrition resistance that exceeded the baseline attrition resistance established for commercial FCC catalysts. All of the fluid-bed catalyst formulations also had some water gas shift activity. Considering that the fluid-bed catalyst formulations were first attempts at preparing a fluid-bed catalyst formulation and more emphasis was placed on achieving attrition resistance, and that the competition was commercial catalysts with decades of R&D aimed at optimizing their performance, our expectation was to achieve a reasonable amount of WGS activity. The HTS catalyst formulation was a huge success with both a high attrition resistance and WGS activity that was comparable with a commercial HTS catalyst. The SGS and LTS catalyst formulations did demonstrate WGS activity and create the opportunity for additional catalyst development focused on optimizing WGS activity.

The preliminary techno-economic analysis demonstrated that the circulating catalyst can act as a large thermal mass allowing the reactor to operate at a lower temperature and achieve more conversion per reactor and that this heat could be recovered as additional high quality steam. The TEA also showed that these technical benefits come with almost a 24% reduction in capital cost.

With the success achieved with the technical feasibility demonstration of ATWGS process, we have created a technology development roadmap that will provide a strategy for continued development of this technology. This roadmap is outlined in **Table 13**.

Table 13. R&D roadmap for ATWGS

Stage	Research objectives
Technical feasibility	• Demonstrate fluidizable WGS catalysts that possess attrition resistance and WGS activity. • Perform preliminary techno-economic analysis to assess cost and efficiency benefits
Catalyst development	• Optimize fluidizable WGS catalysts for attrition resistance and WGS activity • Complete lab-scale testing of catalyst performance using simulated syngas mixtures at typical operating conditions to determine catalytic performance (activity, stability and selectivity) • Update preliminary TEA to assess cost and efficiency benefits
Pilot-plant testing	• Demonstrate performance of a fluidized-bed catalyst in a pilot-scale transport reactor with a solids cooler using real coal-derived syngas.
Demonstration	• Demonstrate expected commercial performance in demonstration-scale equipment with real commercial syngas and a commercially produced catalyst.

7. Coal-to-Fisher-Tropsch Liquids

DOE/NETL has completed and/or funded multiple system studies for the conversion of coal and natural gas into FT products. The key aspects of these different studies are summarized in **Table 14**.

Table 14. Summary of the key aspects of the DOE/NETL system studies FT synthesis (include a landscape table)

Study	Technology components	Comments
Baseline Technical and Economic Assessment of a Commercial Scale Fischer-Tropsch Liquids Facility, April 9, 2007, DOE/NETL-2007/1260" (CTL Study 1)	ConocoPhillips E-Gas gasifier utilizing Illinois #6 coal, Selexol AGR syngas cleanup, Fe-based FT synthesis, and Amine CO ₂ capture within FT synthesis and upgrading process	The Fe-based catalyst performs the necessary WGS in the FT reactor eliminating the need for a separate upstream WGS process and consolidating CO ₂ capture downstream of the FT reactor. Stringent syngas cleaning requirements for the Fe-FT catalyst, but available commercial cleanup technologies cannot decouple removal of CO ₂ and remaining contaminant. Consequently, CO ₂ capture is incorporated at multiple locations within the process.
"Cost and Performance Baseline for Fossil Energy Plants. Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis, October 15, 2014, DOE/NETL-2011/147" (CTL Study 2)	Shell gasifier utilizing Illinois #6 coal, Rectisol® AGR for syngas cleanup, Fe-based FT synthesis, and Amine CO ₂ capture within FT synthesis and upgrading process	The Fe-based catalyst performs the necessary WGS in the FT reactor eliminating the need for a separate upstream WGS process and consolidating CO ₂ capture downstream of the FT reactor. Stringent syngas cleaning requirements for the Fe-FT catalyst, but available commercial cleanup technologies cannot decouple removal of CO ₂ and remaining contaminant. Consequently, CO ₂ capture is incorporated at multiple locations within the process.
Analysis of Natural Gas-to-Liquid Transportation Fuels via Fischer-Tropsch", September 13, 2013, DOE/NETL-2013/1597 (GTL Study)	Natural gas generated syngas with a H ₂ /CO ratio of 2.0–2.2 can be sent directly to the Co-based FT synthesis process with all CO ₂ capture downstream of FT because any contaminants are removed prior to natural gas conversion to syngas.	This report provides extensive details on the FT upgrading process relating to the 30:70 split between naphtha and diesel not provided in the CTL reports.
Bechtel Corporation, Aspen Process Flowsheet Simulation Model of a Battelle Biomass BaseGasification, Fischer-TropschLiquefaction and Combined-Cycle Power Plant. 1998.		This report provides extensive details on the FT upgrading process for both Fe- and Co-based FT synthesis but limited details on the syngas generation and AGR cleanup processes.
Bechtel Corporation, Natural Gas Based Once-Through Fischer-Tropsch Design with Power Coproduction—Topical Report. 1998.		This report provides extensive details on the FT upgrading process for both Fe- and Co-based FT synthesis, but limited details on the syngas generation and AGR cleanup processes.
Bechtel Corporation, Baseline Design/Economics for Advanced Fischer-Tropsch Technology: Final Report. 1998, U.S. Department of Energy, Pittsburgh Energy Technology Center.		This report provides extensive details on the FT upgrading process for both Fe- and Co-based FT synthesis, but limited details on the syngas generation and AGR cleanup processes.

The fundamental challenge in selecting a reference case was none of DOE/NETL's existing cases fully met the project's objective of substantially reducing the cost associated with coal conversion to product with > 90% carbon capture and near zero emissions. DOE/NETL's CTL studies utilize an Fe-based FT process. One of the key advantages of the Fe-based FT catalyst is its ability to effectively use syngas mixtures with H₂ to CO ratios closer to 1, which are typically generated by coal gasification. This ability eliminates the need for any WGS upstream of the FT process. Theoretically, this moves the CO₂ capture process downstream of the FT process. However, the Fe-based FT catalyst also requires extensive syngas cleanup to reduce contaminant poisoning of the Fe-based catalyst and manage catalyst replacement costs. All commercial AGR systems that would be used to meet these syngas cleanup requirements also capture a significant amount of CO₂. As the Fe-based FT process generates significant amounts of CO₂ due to its WGS activity, a second AGR process is needed downstream of FT synthesis. The lower operating pressure of the FT synthesis process also creates more challenges for effective removal of this CO₂ in this second AGR process downstream of FT synthesis.

The Co-based FT process requires both a very clean syngas and very high H₂ to CO ratio for optimal operation. This would add the need for a WGS process prior to FT synthesis, but it would couple the stringent syngas cleanup requirements with bulk CO₂ removal. A CO₂ capture process would still be necessary downstream of the FT reactor to protect the low-temperature hydrocarbon recovery equipment from plugging due to CO₂ freezing. However, this system would be much smaller as it captures ≤ 5% of the total CO₂ captured as CO₂ for the entire plant. This approach does not eliminate the need for two CO₂ removal processes, but it does shift the bulk CO₂ removal to upstream of the FT process where higher syngas pressure facilitates effective removal of both sulfur and CO₂ and optimal use of commercial AGR processes for contaminant removal occurs. Unfortunately, DOE/NETL's existing CTL studies do not include a case with Co-based FT synthesis and the CO-based FT option is only included in GTL studies.

The solution to these challenges was to create a reference case from multiple available DOE/NETL FT studies. In this reference case, a Shell gasifier converts an Illinois #6 coal into a syngas and a Co-based FT synthesis process converts the syngas into FT liquid products. The raw syngas from the gasifier is shifted to achieve the H₂ to CO ratio of 2 and a Rectisol® AGR is used to clean the syngas sufficiently for the Co-based FT catalyst. This reference case is a combination of "Cost and Performance Baseline for Fossil Energy Plants. The key DOE/NETL studies used to build this case include *Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis, October 15, 2014, DOE/NETL-2011/147*" (CTL Study 2) and *Analysis of Natural Gas-to-Liquid Transportation Fuels via Fischer-Tropsch*", September 13, 2013, DOE/NETL-2013/1597 (GTL Study). The DOE/NETL sponsored Bechtel studies were also used to model the FT process as these studies are the basis for the subsequent DOE/NETL studies and provided better documentation of engineering and design details.

In addition to the reference case, two additional advanced CFTL plant configurations were evaluated as part of this case study. In the first of these cases, RTI's advanced warm syngas cleanup and AACRP were used in place of the Rectisol® AGR. In the second advanced plant configuration, the AR advanced gasification system and RTI's advanced warm syngas cleanup system and AACRP were used to generate a syngas from Illinois #6 coal and create a syngas with a H₂ to CO ratio of about 2 and low enough contaminant concentration to be suitable for the Co-based FT synthesis and upgrading process. **Table 15** provides the three specific technology combinations used in this CFTL case study.

Table 15. Case study matrix for coal-to-FT-liquids production with CO₂ capture

	Case 5A ¹	Case 5B	Case 5C
Gasification technology			
Shell gasification	✓	✓	
AR gasification technology			✓
Gas cleanup			
Rectisol® for CO ₂ and sulfur removal ²	✓		

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	Case 5A ¹	Case 5B	Case 5C
RTI WDP with activated amines CO ₂ recovery process (AACRP)		✓	✓
Water gas shift			
Sour shift	✓		
Advanced fixed bed water-gas shift		✓	✓
FT synthesis			
Conventional Co-based FT technology ³	✓	✓	✓
CO ₂ drying and compression (to 2,215 psia)	✓	✓	✓

¹ Case 5A is the reference case. Gasifier selection depends on selected DOE NETL baseline studies.

² Rectisol® removes sulfur and CO₂. Selection and any additional trace contaminant cleanup technologies needed will be as defined by DOE NETL baseline studies.

³ Baseline FT technology will be established based on DOE NETL guidelines/recommendations.

The same factored estimate approach that was used for the IGCC and CTM study was also applied for this study. A key difference was RTI performed the techno-economic analysis for this case study.

RTI developed a conceptual design of the Case 5A reference plant with two-stage Rectisol®-based AGR for H₂S and CO₂ removal, and a slurry bubble column reactor with a Co-based catalyst for FT synthesis. Individual models for the different processes were simulated and validated against published DOE baseline studies. The syngas generation using the Shell gasifier with Illinois #6 coal was validated against data from the Case B1B from “*Cost and Performance Baseline for Fossil Energy Plants, Volume 1b: Bituminous Coal (IGCC) to Electricity, Revision 2b*” DOE/NETL-2015/172, 2015 (2015 DOE baseline study). For FT synthesis, the report, “*Baseline Design/economics for Advanced Fischer-Tropsch Technology: Topical Report VI: Natural Gas Fischer-Tropsch Case: Volume II: Plant Design and Aspen Process Simulation Model*.” 1996, authored by Bechtel under earlier DOE sponsored work on GTL studies was used. The DOE-sponsored Bechtel studies provided details on the FT product distribution from the FT reactor and product upgrading processes. The information from the Co-based FT synthesis case from this report was used in this work. The RTI model for heat recovery steam generator (HRSG) and steam turbine (ST) cycle was validated against published data from Case B1B from the 2015 DOE baseline study. Data from Case B1B was used to determine the steam turbine efficiencies to match the total power and temperature of the steam exiting the turbines. This model was used to determine the steam turbine power in this study. All these validation studies provided confidence that the results from this work are within a reasonable range of accuracy.

In spite of these efforts to fully validate the models, a number of discrepancies came up. The ST power generation for the reference case (Case 5A) is 400 MW. This is higher than the ST power generation in the base case from the DOE CTL study, which is 315 MW. The two cases differ in the extent of the shift reaction prior to the FT synthesis and the choice of the FT synthesis catalyst. The DOE base case uses the unshifted syngas out of the gasifier at a H₂/CO ratio of 0.73 and uses a Fe-catalyst whereas Case 5A shifts the syngas to achieve H₂/CO ratio of 2.0–2.1, and uses the Co-catalyst. The differences in the power generation reflects the improved thermal efficiency resulting from heat recovery through HP steam production in the water gas shift area in Case 5A compared to generation of medium pressure 360 psig steam in the FT area in the DOE CTL study base case.

The variable OPEX costs are dominated by the costs for the FT catalyst, accounting for 18–19% of the total variable operating costs. The initial fill for the FT catalyst was scaled from the Bechtel study using a factored scaling approach. The total initial fill determined was found to be roughly 4 times higher than the catalyst fill reported in the DOE CTL and GTL studies. As this factor was the same for all three CFTL plants in this case study, this fact will not impact any trends and conclusions drawn for this case study. However, this fact will need to be considered when comparing performance results with other DOE

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CTL and GTL studies. The annual consumption rate for the catalyst used in this study was the 2.74 years used in all other DOE studies for the Co-based FT catalyst.

The cost of the FT catalyst was also assumed to be \$20/lb. This is similar to other Co-based catalysts. Both the DOE CTL and the GTL report use a cost of \$7.15/lb for the FT catalyst, even though the CTL study uses a Fe-based catalyst and the GTL study uses a Co-catalyst.

Table 16 shows a summary comparison of the CAPEX, OPEX, FT liquids production, plant power import/export, auxiliary power consumption, FT-products RSP, and cost of CO₂ captured/avoided for the three cases. More details about the process, performance and cost breakdown for each case is provided in the comprehensive TEA report provided as **Appendix C**.

Table 16. CFTL results summary

Case	Case 5A		Case 5B		Case 5C	
Gasifier	Shell		AR		AR	
Sulfur and CO ₂ removal	Dual Stage Rectisol®		RTI WDP/AACRP		RTI WDP/AACRP	
CAPEX, \$MM						
Total installed cost (TIC)	\$3,882		\$2,3734		\$3,220	
Total plant cost (TPC)	\$5405		\$5,187		\$4,577	
Total overnight cost (TOC)	\$6771		\$6,530		\$5,797	
OPEX, \$MM/yr (100% CF basis)						
Fixed operating cost (OC _{fix})	\$205.8		\$198.9		\$179.8	
Variable operating cost, less fuel (OC _{var})	\$176.1		\$190.0		\$179.3	
Coal feedstock (OC _{Coal})	\$526.0		\$526.0		\$526.0	
Power import/(export) (OC _{Power})	(\$21.7)		(\$36.9)		(\$30.7)	
Total OPEX	886.1		\$878.0		\$854.37	
Plant Output						
Crude FT Diesel Product, bbl/day	34,992		34,994		36,115	
Crude FT Naphtha Product, bbl/day	15,008		15,006		15,486	
Net Power Output, MWe	41.3		70.2		58.4	
Power Generation, MWe						
Gas turbine	122.5		101.5		104.9	
Steam turbine	403.2		448.6		414.2	
Syngas turboexpander turbine	7.3		10.0		11.4	
Total gross power	533.0		560.0		530.4	
Net power generation	41.3		70.2		58.4	
Common factors used for RSP calculation						
Capacity charge factor (CCF)	0.170	0.218	0.170	0.218	0.170	0.218
Capacity factor (CF)	0.9		0.9		0.9	
CO ₂ Capture						
CO ₂ capture product, tpd	25,116		25,280		24,729	
Carbon captured (as % carbon in fuel)	92.5%		92.8%		92.6%	
Cost of CO ₂ captured ^{1,2,3} (\$/ton)	13.92	15.0	13.44	14.49	13.47	15.04
Required selling price ^{1,2,3}						
RSP FT diesel, \$/bb _{FTD}	134.3	156.0	131.0	152.0	117.3	135.6
RSP naphtha \$/bbl _{FTN}	94.2	109.2	89.9	104.6	80.4	93.1

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Case	Case 5A		Case 5B		Case 5C	
Equivalent Crude oil price,\$/bbl _{ECO}	116.5	135.4	113.7	131.9	101.8	117.7
Equivalent Petroleum Diesel price, \$/bbl _{EPD}	145.7	169.3	142.1	164.9	127.3	147.1
FT diesel RSP breakdown, \$/bbl						
CAPEX	77.02	98.76	74.27	95.25	63.89	81.93
Fixed O&M	13.77	13.77	13.31	13.31	11.65	11.65
Variable O&M	10.60	10.60	11.44	11.44	10.46	10.46
Fuel	31.67	31.67	31.67	31.67	30.69	30.69
Power generation	-1.31	-1.31	-2.22	-2.22	-1.79	-1.79
CO ₂ T&S	2.50	2.5	2.52	2.52	2.39	2.39
Total	134.25	156.00	131.00	151.97	117.29	135.33

¹ Capacity factor assumed to be 90%.

² Values shown are for two financial structures. The first (lower value) is based on the loan guarantee finance structure. The second (higher value) is based on the commercial fuels finance structure.

³ Includes CO₂ T&S.

Figure 22 illustrates the breakdown of the cost of production of FT-diesel. Since the results for the RSP of FT-naphtha will be similar due to the fixed ratio of naphtha and diesel produced, only the FT-diesel results are shown with the same trends and conclusions applying to the FT-naphtha production. It can be seen from Figure 22 that the cases differ most significantly in terms of capital cost on a \$/bbl basis.

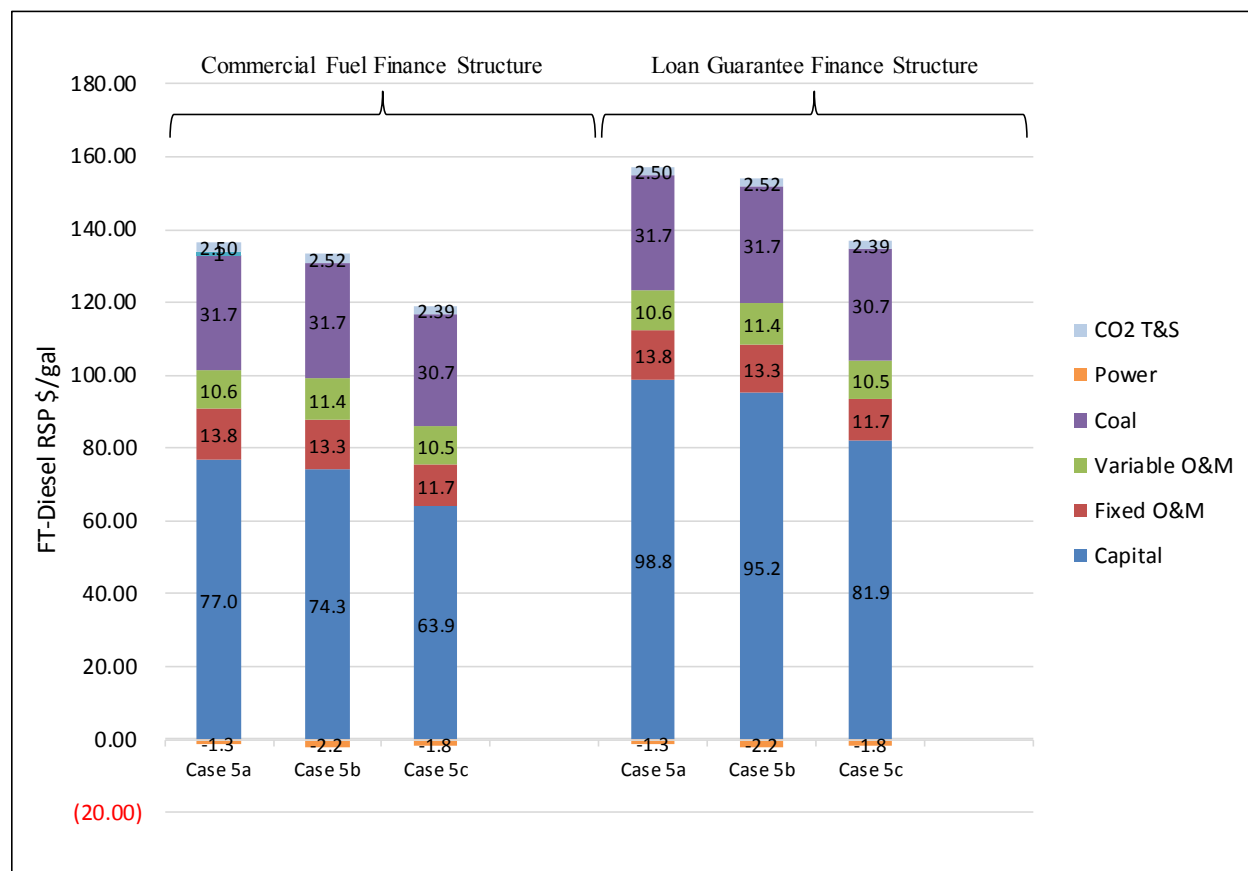


Figure 22. Components of FT diesel

This breakdown by component of the FT-diesel RSP for the different cases illustrates the relative contribution of each component. In all cases, the capital cost accounts for 54–63% of the FT-diesel RSP. Both the AR gasification and RTI advanced warm syngas cleanup technologies significantly reduce capital cost and FT diesel RSP. The AR gasification system reduces the capital cost of the overall gasification block (including the dry feed system) by 35% and results in reduction in TPC of the entire CFTL plant by about \$610 MM or a 11.5% decrease. The RTI advanced warm syngas cleanup system reduces capital costs of the overall gas cleanup block (including AGR, SRU, LTGC, and WGS) by 24% and results in reduction in TPC of the entire CFTL plant by \$211 MM or about 4%. These capital cost reductions result in a 10% and 3% decrease in the FT diesel RSP for the AR gasification and RTI advanced warm syngas cleanup technologies respectively, for both financial structures.

In Case 5A, the CTL plant has a total of 9 Shell gasifiers with 8 in continuous operation and one spare. Based on the coal throughput, the size of these Shell gasifiers is about 2,000 tpd. Based on the assumption that the reference case should reflect actual commercially operating plants, this gasifier selection is practical. However, this study focuses on evaluating emerging technologies which are expected to be demonstrated commercially in the next 5 years. Because of this approach, the AR gasification system included a 3,600 tpd coal gasifier. Recently Shell announced that they are constructing a CTL plant in China with Shell gasifiers with a 3,200 tpd coal intake capacity. If Shell's larger gasifier had been used in the reference case, a total of 6 continuously operating gasifiers would have been required, which is identical to the number of AR gasifiers used in this study. There is no doubt that use of the AR gasification system would still result in a substantial reduction in capital cost (because the case studies in this project have consistently shown that AR gasifiers cost substantially less than equivalent-sized Shell gasifiers), but the total magnitude of any cost reduction would be somewhat smaller.

The gas turbine (GT) power generated by combusting the FT fuel gas is quite different for the three cases. Case 5A produces 122.5 MW of power in the GT. Cases 5B and 5C produce 101.5 MW and 105 MW of GT power, respectively. This difference is due to the reduced hydrogen content in the FT fuel gas due to the hydrogen that is diverted for reducing the SO₂ stream in the DSRP for Cases 5B and 5C. Case 5C uses the AR gasifier which has a higher cold gas efficiency. The benefit from the increased efficiency is utilized by producing excess FT liquids rather than excess power, since the FT liquid is a higher value product.

The ST power produced in the three cases are also quite different. The base case, Case 5A, produces 403 MW of power, compared to 449 MW in Case 5B and 414 MW in Case 5C. Compared to Case 5A, ST power in Case 5B is higher due to the net excess steam production in Case 5B. Case 5B uses the RTI WDP for warm-gas desulfurization. As such, the syngas in Case 5B, instead of being sent to the scrubber, is sent to the RTI WDP process at 750°F, and the gas exiting the RTI WDP is used to raise additional high pressure steam. Also, the AFWGS process uses less steam compared to the sour water gas shift in Case 5A. Both choices result in a larger amount of HP steam that is available for power production in the ST.

Compared to case 5B, Case 5C results in a lower ST power production, even though both processes use the RTI WDP process and the advanced HT shift. This difference in ST power results from the quenching of the syngas that occurs in the AR gasifier compared to the cooling of the syngas in the Shell gasifier by raising HP steam in the radiant and duct coolers of the Shell gasifier.

There is also 7–11 MW of power that is generated by the expansion of the syngas before sending it to the FT synthesis reactor.

Case 5A results in 41 MW of net power that is exported. Due to improved steam generation from process heat recovery and reduced steam consumption in the AFWGS, Case 5B and Case 5C result in 70 MW and 58 MW of net export power, respectively.

The reference case (Case 5A) and thus comparison Cases 5B and 5C were based on use of Illinois #6 coal, which has a high chlorine content (0.29 wt%) and produces a raw syngas with 900 ppmv of HCl. Since high levels of HCl in the syngas are detrimental to all downstream chemical processes, this required a process for removal of the HCl. In Case 5A, the chlorine is removed in the syngas scrubber and the

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chlorine is treated as part of the sour water processing. Case 5B and Case 5C use a warm-temperature non-regenerable HCl guard bed process to remove the HCl from the syngas upstream of the RTI WDP process at high temperature, which improves the overall thermal efficiency of the process as described previously.

Although the technical aspects of the HCl removal could be dealt with, effective costing of the HCl removal was more challenging. Sufficient details were not provided in the DOE reference cases to isolate the chlorine treatment cost from the overall sour gas treatment costs. As all three cases had sour water processing, the factored estimating approach used included the chlorine treatment cost, which was required in Case 5A, but was not needed in Cases 5B and 5C where the chlorine was captured by a warm-temperature non-regenerable HCl sorbent. Replacement of this HCl guard bed sorbent is about \$7MM in the annual operating costs.

If the variable cost of chlorine treatment in the sour water processing is assumed to be equivalent to the variable cost for replacement of the HCl sorbent, the cost of chlorine removal is about 5% of the total variable annual operating cost. Since the variable operating expenses only contribute to ~10.5% of the RSP, the cost of the additional guard bed contributes to only ~0.5% of the total RSP. Thus, the overestimated cost of chlorine removal for Cases 5B and 5C is about 5% of total annual variable costs and 0.5% of diesel RSP. However, a more detailed analysis of these costs should be conducted to effectively understand the cost of chlorine removal for high chlorine coals for these advanced technologies. For most other coals, which do not contain such high levels of chlorine, the use of HCl guard beds would not be required and their resultant capital and operating costs would thus be avoided, which would increase the relative economic benefits for Cases 5B and 5C reported above.

A key criterion for the CFTL plants in this study was to achieve near-zero levels of emissions. Because of the sensitivity of the cobalt-based FT catalyst used in the FT synthesis process, the gas phase sulfur emissions were effectively reduced to zero. Although other contaminants were not tracked as meticulously as sulfur, their gas phase emissions were also essentially reduced to zero because of the syngas cleaning required for the cobalt-based FT catalyst.

Because the byproduct hydrocarbons from the FT synthesis and upgrading processes are used for fuel gas and to generate electricity, there are some net CO₂ gas phase emissions. However, all three cases result in between 92% and 93% carbon capture, which is because CO₂ capture is an inherent part of the FT process requirements. Consequently, the cost of CO₂ captured ranges from \$13.40–\$13.90 per barrel of CO₂ for all three cases for the loan guarantee finance structure, because the only required additional storage and sequestration cost comes from compression of the CO₂ captured to 2215 psia for sequestration. Interestingly, the CO₂ footprint for the diesel and naphtha products is about 147 lb of CO₂/bbl for Case 5A and 143 lb CO₂/bbl for Cases 5B and 5C. With all the net CO₂ emissions being assigned to the FT products, the electricity being generated has a zero-carbon footprint. Because the electricity can be viewed as an inherent byproduct of the CFTL plant, market forces will attempt to drive down the price of this electricity. Because this electricity has essentially no carbon footprint, it is possible that it could be negotiated to command a premium price in the proper regulatory environment.

Therefore, the CFTL plant achieves lower gas phase emissions than either the IGCC or CTM plants also studied under this project. But even for these CFTL plants, the advanced technologies, and, in particular, RTI's warm syngas cleanup technologies result in emission reductions.

Many of the key findings from the sensitivity studies match the results seen in the previous IGCC and CTM case studies. The benefits associated with AR's advanced gasification and RTI's advanced warm syngas cleanup technologies were still present even if the equipment cost for these technologies was increased by 30%. The selling price for any excess electricity generated and sold had only a small (< \$10/ton) impact on the diesel RSP for a COE between 0 and \$120/MWh. Finally, the greatest benefits were seen for the case integrating both AR's advanced gasification and RTI's advanced warm syngas cleanup technologies.

The substantial benefits from integration of AR's gasification and RTI's advanced syngas cleanup technologies into a CFTL plant, which include significant capital cost reductions, lower emissions, and improved energy efficiency, will generate market forces that will drive the implementation of the AR and

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RTI technologies to realize these benefits. All these factors indicate that when ready for commercial deployment, these technologies will be favored over existing commercial technologies.

8. Conclusion

The overall objective of this project was to assess the potential to achieve a substantial reduction in the production cost of H₂-rich syngas via gasification with near-zero emissions due to cumulative and synergistic improvements realized when multiple advanced technologies are incorporated into the overall process.

Our selection of advanced emerging technologies included the AR advanced gasification technology and the RTI advanced warm syngas cleanup technologies. Within the process blocks that make up RTI's advanced warm syngas cleanup, we have included not only the WDP and DSRP processes, but also included the AFWGS process. In work conducted in parallel with this project, these technologies have been undergoing pilot and pre-commercial demonstration testing. AR's advanced gasification technology was acquired by GTI and is now being offered commercially as R-GAS™. Similarly, RTI has partnered with Casale SA to begin offering commercial licenses for RTI's WDP. Thus, as these technologies begin to penetrate the market, the results from this project will be available to demonstrate the potential technical and economic performance of these technologies.

To fully evaluate the potential benefits of these technologies for producing H₂-rich syngas via gasification, we completed TEAs for three different cases studies, which included:

- IGCC with > 90% carbon capture,
- CTM with > 90% carbon capture, and
- CFTL with > 90% carbon capture.

The case studies consisted of using a reference case completed by DOE/NETL in one of their available system studies based on current commercial technology, which was then compared with different plant configurations incorporating the advanced technology processes in different combinations. Because of the goal to use these TEAs for comparison to identify and quantify benefits, these TEAs were conducted meticulously according to DOE/NETL's QGESS. For the IGCC and CTM case studies, the case study consisted of the reference cases and three plant configurations with the advanced technologies. Two of the plant configurations included either AR's or RTI's advanced technology package. The final case study incorporated both advanced technology packages. The CFTL case study included the reference case and two advanced plant configurations. One of the plant configurations had both AR's and RTI's advanced technology packages. The other had just RTI's advanced syngas cleanup system.

For all three of the case studies, the plant configuration with the best overall performance was always the integrated advanced plant with both AR's and RTI's advanced technology packages. The benefits of these integrated plants included capital and operating cost reductions, higher efficiency or extra power generation, and reductions in emissions. **Table 17** summarizes these benefits for the three plants from the case studies that incorporated both AR's and RTI's advanced technology packages.

From the sensitivity analyses completed, the cost benefits remained even when the cost of the advanced processes were increased by 30%. Considering that the cost data for these TEAs was derived from actual costs for construction of pilot or pre-commercial demonstration systems, the relative error in cost estimation for the advanced technologies should be less than 30% supporting the magnitude of cost benefits identified in Table 17.

Throughout the case studies there were many examples of synergistic benefits that came with the integration of the advanced technologies. Examples included the benefits from the higher operating pressure for AR's gasification system with Rectisol®, reduced steam consumption with RTI's AFWGS, and even the ability to reduce contaminant concentrations to comparable to post Rectisol® concentrations at roughly 70% of the capital cost from decoupling the sulfur and CO₂ removal with WDP and AACRP.

Table 17. Summary of benefits identified in case studies

	IGCC	CTM	CFTL
Product cost reductions	18 mills/kWh	\$77–93/ton	\$17–21/bbl ¹
Capital cost reduction	18%	23%	15%
Operating cost reduction	12%	13%	3.6%
Efficiency Improvement	1.4%		
Surplus power generated		28 MW ²	41%
Emission reduction			
Sulfur	75–79%	99%	
CO ₂ emissions		13–15%	4 lb CO ₂ /bbl ³

¹ This cost reduction applies for both the diesel and naphtha products produced.

² The reference case consumed approximately 1 MW of external power.

³ The CFTL cases inherently remove the CO₂ to optimize the FT process, but the advanced technologies manage to reduce net CO₂ emission per unit of product produced.

The technical feasibility analysis for RTI's ATWGS demonstrated that this process can reduce capital costs compared to a conventional fixed-bed WGS system by about 24%. Evaluation of actual catalyst samples was successful at producing attrition resistant formulations for HTS, LTS and SGS, but only HTS-1, the HTS formulation, had activity that was approximately equivalent to a commercial catalyst. Because of these positive results, R&D continues with the objective of moving towards a pilot plant demonstration of this technology.

By far, the most positive impact from this project is that as these technologies enter the market, market forces will drive the implementation of these advanced technologies to effectively realize their associated cost benefits. Because RTI's advanced warm syngas cleanup process also results in emission reductions in addition to cost benefits, market forces will also result in the implementation of cleaner technologies. **All these factors indicate that now that these technologies are ready for commercial deployment, they will be favored over existing commercial technologies.**

9. Acronyms

°C	Degrees Celsius
°F	Degree Fahrenheit
\$/bbl	Dollars per barrel
\$/gal	Dollars per gallon
\$/kW	Dollars per kilowatt
\$/MWh	Dollars per megawatt hour
\$/tonne	Dollars per ton
\$MM/yr	Million dollars per year
ΔH	Enthalpy
μm	micron
H ⁻¹	per hour (unit of measurement for space velocity)
3D	Three dimensional
AACE	Advancement of Cost Engineering International
AACRP	Activated amine carbon dioxide recovery process
AFWGS	Advanced fixed-bed water gas shift
AGR	Acid gas removal
Al ₂ O ₃	Alumina
AR	Aerojet Rocketdyne
Ar	Argon
ATWGS	Advanced transport reactor-based water gas shift
BOP	Balance of plant
Bbl/d	Barrel per day
CF	Capacity factor
Capex	Capital costs
CO ₂	Carbon dioxide
CO	Carbon monoxide
COS	Carbonyl sulfide
Cr ₂ O ₃	Chromium oxide
Co	Cobalt
CoO	Cobalt oxide
CH ₄	Methane
CFD	Computational fluid dynamics
CGE	Compact gasification system
CuO	Copper oxide
COE	Cost of electricity
CFTL	Coal-to-Fischer-Tropsch-liquids
CTL	Coal-to-Liquids
CTM	Coal-to-methanol
DI	Davison index
DOE	U.S. Department of Energy
DOE/NETL	U.S. Department of Energy National Energy Technology Laboratory
DSRP	Direct sulfur recovery process
DSP	Dry solids pump
EERC	Energy and Environmental Research Center
Fe	Iron
Fe ₂ O ₃	Iron oxide
FT	Fischer-Tropsch
FCC	Fluid catalytic cracking

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GC	Gas chromatograph
GTI	Gas Technology Institute
GT	Gas turbine
GE	General Electric
GWh/y	Gigawatt hours per year
g/cm ³	Gram per centimeter cubed
H ₂	Hydrogen
H ₂ O	Water
H ₂ S	Hydrogen sulfide
HCl	Hydrochloric acid
HMB	Heat and mass balance
HRSG	Heat recovery steam generator
HHV	Higher heating value
HTS	High temperature shift
HTS-1	RTI's high temperature shift fluid-bed catalyst formulation
In	Inch
ID	Internal diameter
Lb/MMh _{gross}	Pounds per million hours gross
Lb/MMBtu	Pounds per million British Thermal units
LHV	Lower heating value
LTGC	Low temperature gas cooling
LTS	Low temperature shift
LTS-1	RTI's low temperature shift fluid-bed catalyst formulation
MgO	Magnesium oxide
MW	Megawatt
MW _e	Megawatt of electricity
Mtpd	metric tons per day
MM	Million
MMBtu/d	Million British thermal units per day
Mill/kWh	Mills per kilowatt hour
MoO ₃	Molybdenum oxide
NGCC	Natural gas combined cycle
OC _{coal}	Operating cost associated with coal usage
OC _{fix}	Fixed operating cost
OC _{fuel}	Fuel related operating cost
OC _{NG}	Operating cost associated with natural gas usage
OC _{POWER}	Operating costs associated with power
OC _{var}	Variable operating cost
Opex	Operating costs
O&M	Operating and maintenance
PCC	Pulverized coal combustion
Ppmv	parts per million by volume
Psi	Pounds per square inch
Psig	pounds force per square inch guage
Psia	pounds force per square inch absolute
PRB	Powder river basin
QGESS	Quality guidelines for energy system studies
R&D	Research and development
R-GAS TM	GTI's marketing name for Aerojet Rocketdyne's advanced gasification technologies
RSP	Required selling price
RTI	RTI International

Benefits of Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production

SGS	Sour gas shift
SGS-1	RTI's sour gas shift fluid-bed catalyst formulation
ST	Steam turbine
SNG	Substitute natural gas
S	Sulfur
SRU	Sulfur recovery unit
SO ₂	Sulfur dioxide
TEA	Techno-economic analysis
TEC	Tampa Electric Company
TIC	Total installed cost
TOC	Total overnight cost
TPC	Total plant cost
tpd	Ton per day
TS&M	Transportation storage and maintenance
vol%	volume percent
WDP	Warm desulfurization process
WGS	Water gas shift
ZnO	Zinc oxide

10. Appendix A

Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production

Task 2: Techno-Economic Analysis for IGCC Application

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Submitted for

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PREFACE

The primary factor that motivated the preparation of the techno-economic analyses completed in this study was new technical and cost data that became available based on pilot plant testing completed on Aerojet Rocketdyne's dry solids pump and gasifier and construction and operation of RTI's 50 MWe pre-commercial system at Tampa Electric's Polk Power Station in Tampa, FL. Over the course of completing these studies, a number of the hypotheses that we had proposed were confirmed. One of these was that the integration of multiple advanced technologies can result in synergistic improvements in capital and operating costs and efficiency. It was through integration of RTI's Warm Gas Desulfurization Process (WDP) for sulfur removal and an activated amine process for CO₂ removal that we recognized not only the improvement in sulfur and CO₂ removal that could be achieved, but also the significant benefits in lower capital and operating costs when our technology was integrated with conventional commercial CO₂ removal technologies. The other hypothesis that we validated through these studies was that true integration does not stop at the boundary limits of the specific technology or technologies being studied, but readjusts the entire system or plant for optimal performance. Through such system-wide optimization studies, each technology is really compared at its full potential reflecting the way a technology or more specifically an emerging technology will eventually be applied commercially. We believe that this extra effort on the team's part and the involvement of the technology developers sets this study apart and provides a candid and real look at technologies that will be entering the market place within the next few years.

Although we are very excited about sharing the results from this study, we recognize that there is a significant gap between this study and some of the previous studies of our warm syngas cleanup technologies. Some of this gap was created by the improved capital and operating costs available based on construction and operation of the pre-commercial test system that were not available for the previous study. However, the most significant source of the gap was that this study included a requirement for 90% carbon capture. This 90% carbon capture requirement results in significant addition of equipment, capital and operating cost for both baseline and advanced technology options as well as thermal efficiency losses and increased parasitic power load. Furthermore, current commercial technologies remove both sulfur and CO₂ in a single process. Thus, the only means of comparing the technologies was by assessing the total syngas cleanup system equipment block. But, the amount of CO₂ captured is roughly 60 times the amount of sulfur captured, which means the total syngas cleanup system is significantly larger for a system which includes 90% carbon capture and the relative contribution of the sulfur removal system becomes much smaller. As a consequence, any system-wide benefits introduced by our desulfurization technology will be diluted and will appear relatively smaller in this study when compared with the relative benefits cited in the previous study that did not require 90% carbon capture.

Unfortunately, the scope of work for these studies did not allow us to use the new information available from the WDP pre-commercial test unit to directly update the previous study for IGCC without carbon capture. However to help potential end users of our desulfurization technology, RTI with its own funding conducted a techno-economic analysis that compared our WDP technology with a conventional desulfurization technology (Selexol™) in an Integrated

Gasification Combined Cycle (IGCC) application that did not include carbon capture, but did include the updated costs from the pre-commercial demonstration unit. Figure P-1 shows that in a comparison of the capital cost of Selexol™ and WDP desulfurization systems for an IGCC system generating about 600 MWe, RTI's WDP system results in a 20% reduction in capital costs. Figure P-2 shows a comparison of the operating costs for these same IGCC systems. RTI's WDP system also significantly reduces the overall operating costs. The main benefit comes from a significant increase in the net amount of high-quality steam that is generated because the syngas does not have to be cooled down for the desulfurization process. These results show that even as improved and technical information associated with WDP becomes available, the economic benefits for RTI's WDP over conventional desulfurization technologies remain strong. With the completion of testing with the 50 MWe pre-commercial test unit, we look forward to successful commercial deployment of WDP in the next few years.

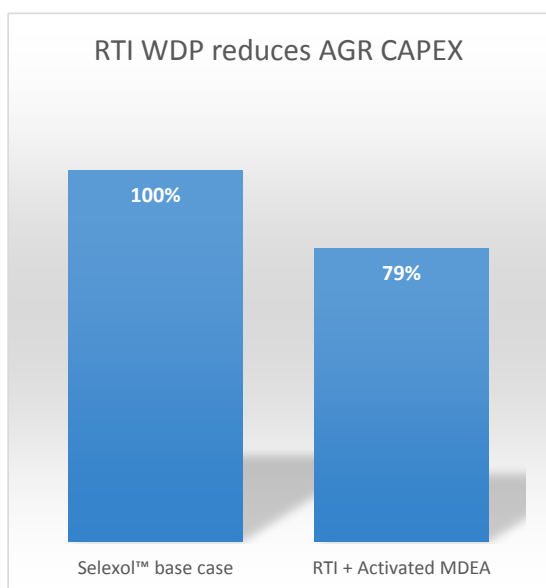


Figure P-1. Comparison of CAPEX for IGCC without carbon capture

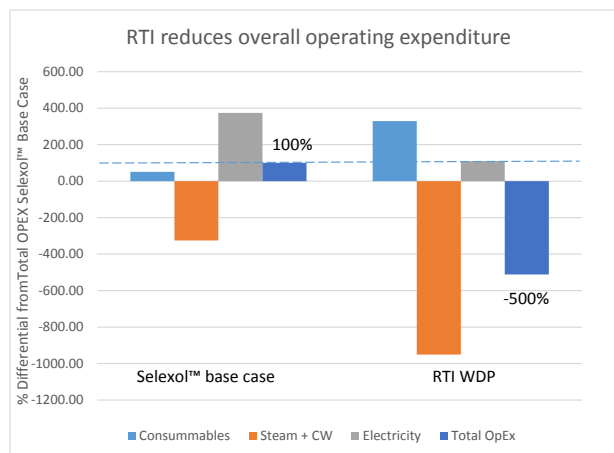


Figure P-2. Comparison of OPEX for IGCC without carbon capture

October 2016

Brian Turk and David Denton (RTI)

ABSTRACT

This study evaluated the techno-economic benefits of integrating Aerojet Rocketdyne's (AR's) advance gasification technologies, (currently being offered by the Gas Technology Institute as R-GAS™) with RTI's advanced warm syngas cleanup technologies for hydrogen-rich syngas production for Integrated Gasification Combined Cycle (IGCC) applications with >90% carbon capture. The reference case was a Shell IGCC power plant with dual stage Selexol™ acid gas removal system (Case S1b from DOE/NETL's "Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity, May 2011"). In addition to the reference case, three additional cases were developed. In the first study, both advanced technologies were integrated into the IGCC plant. In the final two of these studies, AR's and RTI's advanced technologies were integrated into the IGCC plant individually.

In comparison with this reference case, the IGCC case with both AR's and RTI's advanced technologies offered the following benefits:

- ~ 1.4% net efficiency improvement,
- ~18% reduction in capital cost per kW generated,
- ~12% reduction in the operating cost per kW generated, and
- ~ 75-79% reduction in sulfur emissions.

Furthermore, the sensitivity analysis completed shows that the benefits in cost of electricity (COE) for AR's gasification and RTI's WDP technologies still exist even for a 30% increase in the estimated total plant costs for these technologies.

The fact that the fully integrated IGCC case offers both economic benefits and reduces both sulfur and CO₂ emissions creates a unique condition where market forces will drive the implementation of such cleaner technologies. The significant economic and efficiency benefits of the fully integrated IGCC case will favor commercial deployment of these advanced technologies over existing commercial technologies.

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EXECUTIVE SUMMARY

With U.S. Department of Energy National Energy Technology Laboratory (DOE/NETL) funding, Research Triangle Institute (RTI), Aerojet Rocketdyne (AR) and Nexant completed a set of techno-economic analyses to assess the potential benefits associated with integrating the emerging technologies being offered by RTI and AR for reducing the production costs for hydrogen-rich syngas with near-zero emissions from coal gasification for both integrated gasification combined cycle (IGCC) and Coal-to-Liquids (CTL) applications.

RTI's Warm Gas Desulfurization Process and AR's Advanced Gasification System are emerging technologies that will help enable and drive commercial deployment of gasification through reduced costs, increased efficiency and greater contaminant removal within the next five years. AR's compact gasifier is an entrained-flow, dry-feed gasification system that utilizes AR's proprietary dry feed pump. Due to their novel designs, AR's compact gasifier and dry solids pump (DSP) significantly reduce the size and amount of equipment, which is projected to reduce the cost of the gasification system by approximately half.

RTI's advanced syngas cleanup processes consist of the following components: RTI's warm gas desulfurization process (WDP), RTI's direct sulfur recovery process (DSRP), and RTI's advanced fixed-bed water gas shift process (AFWGS). An activated amine carbon dioxide recovery process (AACRP) has also been integrated with RTI's advanced warm syngas cleanup process, because through this integration contaminant removal equivalent to that of Rectisol[®] can be achieved at a fraction of the cost. With RTI's 50 MWe demonstration plant currently demonstrating the technology at Tampa Electric Company's Polk Power Station, RTI advanced syngas cleanup technology will be ready for commercial deployment in the next year.

To fully assess the benefits of each advanced technology, a total of four techno-economic analyses, which are shown in Table E-1, were completed which included >90% carbon capture. The first of these techno-economic analyses was a reference case, which was Case S1B from DOE/NETL's "*Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity*, May 2011". This case was selected specifically because of the use of the Shell gasification train for the following reasons:

- The Shell gasifier is a dry feed gasifier, which would offer the best comparison with the dry feed AR gasifier,
- The DOE/NETL baseline studies have demonstrated that an IGCC based on the Shell gasifier had the highest efficiency, and
- The Shell gasification technology is a commercially mature technology.

Based on the four cases shown in Table E-1, Nexant developed heat and mass balances for the four cases. Then applying the assumptions, and methodology established in DOE/NETL's Series of Quality Guidelines for Energy Systems Studies (QGESS), Nexant developed factored cost estimates for each case. Nexant used the reference case to validate and fine tune their techno-economic methodology. Using DOE/NETL's methodology, AR and RTI developed the necessary heat and mass balances, equipment costs, and utility and consumable needs and cost for their advanced technology processes based on their pilot plant and demonstration experience.

The results from these case studies are summarized in Table E-2. Once these cost estimates had been completed, Nexant also conducted sensitivity analyses for capital cost, feedstock cost, IGCC capacity factor, CO₂ sales prices and cost of CO₂ emissions.

Table E-1
Case Study Matrix for IGCC with CO₂ Capture

	Case 2a ¹	Case 2b	Case 2c	Case 2d
Gasification Technology				
Shell Gasification with Lockhopper-Based Feed System	✓			✓
AR Compact Gasifier with Dry Solids Pump (DSP) Feed System		✓	✓	
Gas Cleanup ²				
Two-Stage Selexol™ for CO ₂ and Sulfur Removal	✓		✓	
RTI WDP with AACRP		✓		✓
Water Gas Shift				
Sour Shift	✓		✓	
AFWGS		✓		✓
GE 7F Advanced Gas Turbine	✓	✓	✓	✓
CO ₂ Drying and Compression (to 2,215 psia)	✓	✓	✓	✓

¹ Reference case based on Nexant's benchmark simulation of the NETL Report 1399 Case S1B

² Selexol™ removes H₂S and CO₂. Additional trace contaminant cleanup technologies will be included as defined by DOE/NETL baseline studies

Table E-2 clearly quantifies the benefits associated with the integration of AR advanced gasification and RTI's advanced syngas cleanup technologies in an IGCC with carbon capture. Comparison Case 2b, in which these two advanced technologies were integrated achieved the best overall performance of all comparison cases completed in all the following categories:

- Overall plant efficiency: 32.75%
- Capital Cost: \$2,487 MM
- Total operating cost: \$106.2 MM
- COE: 122.0 mills/kWh
- Cost of CO₂ avoided: 51.1 \$/ton CO₂
- SO₂ emissions: 0.0013 lb/MWh_{gross}

From these conclusions, the integration of AR advanced gasification and RTI advanced syngas cleanup processes results in the following benefits when compared with the best available commercial technologies:

- ~ 1.4% net efficiency improvement,
- ~18% reduction in capital cost per kW generated,
- ~12% reduction in the operating cost per kW generated, and
- ~ 75-79% reduction in sulfur emissions.

Table E-2
IGCC Results Summary

Case	Case 2a	Case 2b	Case 2c	Case 2d
IGCC Configuration				
Gasifier	Shell	AR	AR	Shell
Sulfur and CO ₂ Removal	Selexol™	RTI WDP	Selexol™	RTI WDP
CAPEX, \$MM				
Total Installed Cost (TIC)	\$1,521	\$1,242	\$1,309	\$1,444
Total Plant Cost (TPC)	\$2,030	\$1,674	\$1,772	\$1,919
Total Overnight Cost (TOC)	\$2,487	\$2,048	\$2,174	\$2,344
OPEX, \$MM/yr (100% Capacity Factor Basis)				
Fixed Operating Cost (OC _{fix})	\$74.2	\$62.8	\$65.9	\$70.6
Variable Operating Cost, less Fuel (OC _{var})	\$56.9	\$49.1	\$50.5	\$55.2
Fuel (OC _{fuel})	\$50.4	\$48.4	\$48.4	\$50.4
Total OPEX	\$181.5	\$160.2	\$164.8	\$176.2
Power Production, MWe				
Gas Turbine	430.0	429.9	430.2	429.6
Steam Turbine	224.1	211.3	209.3	226.4
Auxiliary Power Consumption	193.4	178.7	177.8	192.2
Net Power Output	460.6	462.5	461.8	463.8
Emissions				
SO ₂ Emissions (lb/MWh _{gross})	0.0063	0.0013	0.0053	0.0014
SO ₂ Emissions (lb/MMBtu)	0.0008	0.0002	0.0007	0.0002
Fuel Rate and Efficiency				
Coal Feed Rate, tpd AR Coal	7,032	6,752	6,752	7,032
Net Efficiency, %	31.32%	32.75%	32.70%	31.53%
Power Generated, MWh/yr (MWh)	4,034,901	4,051,626	4,045,045	4,062,618
TOC CAPEX, \$/kW	5,400	4,428	4,709	5,054
Total OPEX, \$/MWh	44.97	39.55	40.74	43.36
COE, excl CO₂ TS&M, mills/kWh	145.3	122.0	128.3	137.3
COE, incl CO₂ TS&M, mills/kWh	166.2	142.0	148.4	158.1
Cost of CO₂ Avoided excl CO₂ TS&M, \$/ton CO₂	79.8	51.1	59.0	70.0
Cost of CO₂ Avoided incl CO₂ TS&M, \$/ton CO₂	105.1	75.2	83.1	95.0

Because of the significant economic and efficiency benefits, market forces will drive commercial implementation of these technologies to realize these economic benefits. The fact that integration of these two technologies optimizes the benefits will also drive integrated deployment of these technologies. The fact that the RTI advanced syngas cleanup process also reduces sulfur emissions creates a condition where market forces will also drive the implementation of cleaner technologies.

The sensitivity analysis completed shows that the benefits in COE for AR's gasification and RTI's WDP technologies still exists for a 30% increase in the total plant cost for these technologies. All of these factors indicate that when ready for commercial deployment, these technologies will be favored over existing commercial technologies.

Section 1 Introduction

1.1 BACKGROUND

With U.S. Department of Energy National Energy Technology Laboratory (DOE/NETL) funding from Cooperative Agreement DE-FE0012066, Research Triangle Institute (RTI), Aerojet Rocketdyne (AR), and Nexant formed a team to evaluate the cost and performance benefits associated with integration of the advanced gasification technologies being developed by AR and the warm gas desulfurization process being developed by RTI for hydrogen-rich syngas production for Integrated Gasification Combined Cycle (IGCC) and Coal-To-Liquids (CTL) applications with $\geq 90\%$ carbon capture.

RTI's Warm Gas Desulfurization Process and AR's Advanced Gasification System are next generation technologies that will enable commercial deployment of gasification through improved efficiency, greater contaminant removal, and lower costs within the next five years. AR's compact gasifier is an entrained-flow, dry-feed gasification system that utilizes a dry solids pump (DSP) as the coal feed system. Due to its novel design, the gasifier is drastically smaller (90% size reduction) than a conventional gasifier. Used in conjunction with the DSP, which replaces a large and costly lockhopper-based pressurized coal feed system, the overall compact gasification system is projected to cost as little as 50% of that of a conventional gasification system.

RTI's advanced syngas cleanup process consists of the following components: RTI's Warm Gas Desulfurization Process (WDP), RTI's Direct Sulfur Recovery Process (DSRP), and RTI's Advanced Fixed-Bed Water-Gas Shift Process (AFWGS). An activated amine carbon dioxide (CO₂) recovery process (AACRP) is also integrated with RTI's advanced syngas cleanup process to effectively remove CO₂. This combination effectively decouples the separation of the sulfur and CO₂ enabling more efficient removal of these contaminants at lower cost. Because this combination is also able to produce syngas with contaminant concentrations similar to a Rectisol® process, RTI's advanced syngas cleanup process offers the potential for significant reduction in contaminant emissions at lower cost than commercially available technologies. With RTI's 50 MWe demonstration plant project (DE-FE0000489) currently demonstrating the technology at Tampa Electric Company's Polk Power Station, RTI's advanced syngas cleanup technology will be ready for commercial deployment within the next year.

1.2 STUDY OBJECTIVES

The overall objective of this project is to assess the potential of integrating the technologies offered by RTI and AR in order to reduce the capital and production costs for hydrogen-rich syngas with near-zero emissions from coal gasification for both power and CTL productions with carbon capture. To fully evaluate the integration of these advanced technologies, separate studies were completed for hydrogen-rich syngas production for power and CTL applications. This report documents the study completed for power production for an IGCC. Separate reports are being prepared for the CTL production studies.

The specific case studies completed provide a comparison of each advanced technology individually and as an integrated plant with a reference plant using commercially available technologies. All comparison studies conducted for this report capture 90% CO₂ for storage.

Section 2 IGCC Design Basis

2.1 DESIGN REFERENCES

The reference plant design used for this study was selected from “Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity, May 2011, DOE/NETL. 2010/1399” (NETL Report 1399). NETL Report 1399 contains a series of IGCC designs based on various gasifiers. The reference IGCC case used for comparison against the AR gasification and RTI advanced syngas cleanup and AACRP systems was Case S1B. NETL Report 1399, along with the following DOE/NETL’s Series of Quality Guidelines for Energy Systems Studies (QGESS):

- “Specifications for Selected Feedstocks, January 2012, DOE/NETL. 341/011812”,
- “Process Modeling Design Parameters, January 2012, DOE/NETL. 341/081911”,
- “CO₂ Impurity Design Parameters, January 2012, DOE/NETL. 341/011212”,
- “Detailed Coal Specifications, January 2012, DOE/NETL-401/01211”,
- “Cost Estimation Methodology for NETL Assessments of Power Plant Performance, April 2011, DOE/NETL. 2011/1455”,
- “Capital Cost Scaling Methodology, January 2013, DOE/NETL. 341/013113”, and
- “Fuel Prices for Selected Feedstocks in NETL Studies, November 2012, DOE/NETL 341/11212”

contain a comprehensive set of IGCC design bases and assumptions, as well as reference costs and economic evaluation guidelines that were used to complete the techno-economic analyses (TEAs) in this report. DOE/NETL’s report, “Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases, August 2012, DOE/NETL-341/082312” (NETL Report 341/082312), was also used as the reference to develop the updated capital and operating cost estimates in June 2011 dollars.

2.2 CASE CONFIGURATIONS

To identify and determine any synergistic advantages of integrating both of these technologies, a total of four case studies were completed. One of these cases was a reference case, which was Case S1B selected from NETL Report 1399. The most promising case study was anticipated to be an IGCC plant with CO₂ capture integrating both AR’s advanced gasification and RTI’s advanced syngas cleanup process. The final two IGCC cases evaluated utilizing either the AR advanced gasification system or RTI advanced syngas cleanup process on a standalone basis.

The specific technologies included in each of the four IGCC configurations are identified in the IGCC case study matrix shown in Table 2-1.

Table 2-1
Case Study Matrix for IGCC with CO₂ Capture

	Case 2a ¹	Case 2b	Case 2c	Case 2d
Gasification Technology				
Shell Gasification with Lockhopper-Based Feed System	✓			✓
AR Compact Gasifier with Dry Solids Pump (DSP) Feed System		✓	✓	
Gas Cleanup ²				
Two-Stage Selexol™ for CO ₂ and Sulfur Removal	✓		✓	
RTI WDP with AACRP		✓		✓
Water Gas Shift				
Sour Shift	✓		✓	
AFWGS		✓		✓
GE 7F Advanced Gas Turbine	✓	✓	✓	✓
CO ₂ Drying and Compression (to 2,215 psia)	✓	✓	✓	✓

¹ Reference case based on Nexant's benchmark simulation of the NETL Report 1399 Case S1B

² Selexol™ removes H₂S and CO₂. Additional trace contaminant cleanup technologies will be included as defined by DOE/NETL baseline studies

2.2.1 Case 2a: Reference Shell IGCC Power Plant with Selexol™-Based AGR

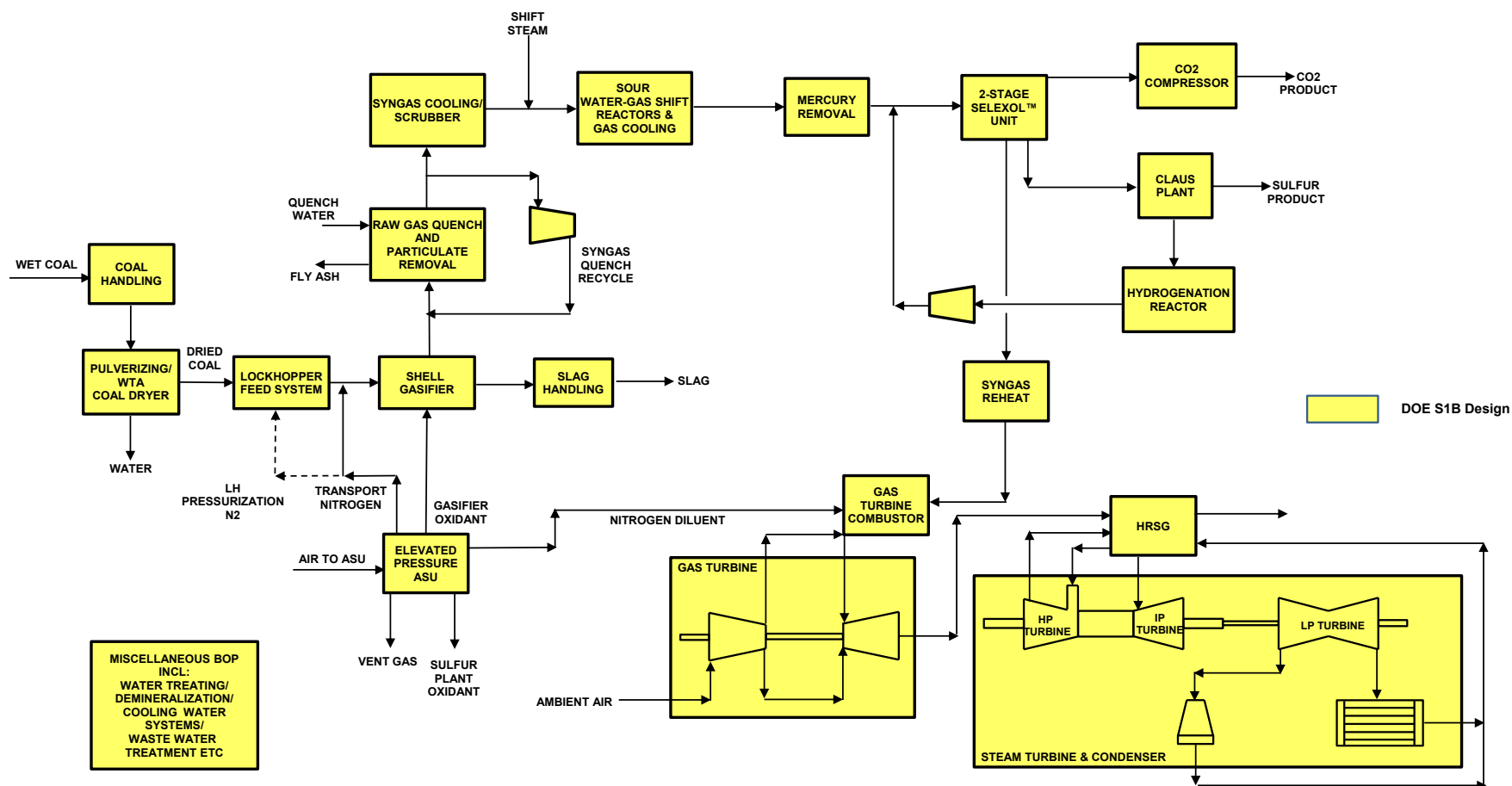
Upon review of the various IGCC cases published in the NETL Report 1399, the Shell gasification-based IGCC case with CO₂ capture utilizing Montana Powder River Basin (PRB) subbituminous coal (Case S1B from NETL Report 1399) was selected as the reference case, for the following reasons:

- The Shell gasifier-based IGCC utilizes a pressurized lock hopper dry feed system. This provides a more direct comparison against AR's DSP feed system, which is part of its overall advanced gasification package. Because the reference and AR advanced gasification systems are based on dry coal feed, a more direct comparison can be drawn between the two, with minimal secondary integration impacts associated with differences in ancillary equipment/subsystems.
- Of all the gasification technologies evaluated in the DOE/NETL Baseline Studies, the Shell gasification-based IGCC has the highest efficiency. If the IGCC utilizing AR/RTI technologies can perform better than the Shell IGCC design in terms of efficiency and cost, it will clearly establish the integrated AR advanced gasification and RTI advanced syngas cleanup system as the best available IGCC technology.
- Shell's gasification technology is considered mature, having already been proven on a commercial scale. Therefore, overall performance and cost can be estimated at a higher confidence level.

The reference Shell gasification-based IGCC case is a coal-fired IGCC plant generating enough hydrogen-rich fuel gas to fill two advanced General Electric (GE) 7F-turbines rated nominally at 215 MWe each, for a total of 430 MWe at the Montana site's elevation, while enabling 90% capture of the carbon in the raw syngas. The power plant is also equipped with a heat recovery steam generator (HRSG) and steam turbine (ST) to generate additional power from waste heat from the flue gas. Adding in the steam turbine power and subtracting auxiliary loads (including

CO₂ capture and compression), the reference IGCC plant's nominal net export capacity is 472 MWe. This reference IGCC plant includes both water gas shift (WGS) and two-stage Selexol™ processes to achieve the required sulfur and CO₂ removal. The reference case has a capacity factor (CF) of 80%. The remainder of the comparison cases were also evaluated at an 80% CF, because this reflects the maximum availability demonstrated by commercial IGCC plants. A simplified reference Case 2a IGCC plant Block Flow Diagram (BFD) is shown in Figure 2-1.

Figure 2-1
Case 2a: Reference Shell Gasifier with Selexol™ AGR IGCC - Simplified BFD



2.2.2 Case 2b: AR Compact Gasifier with RTI WDP IGCC

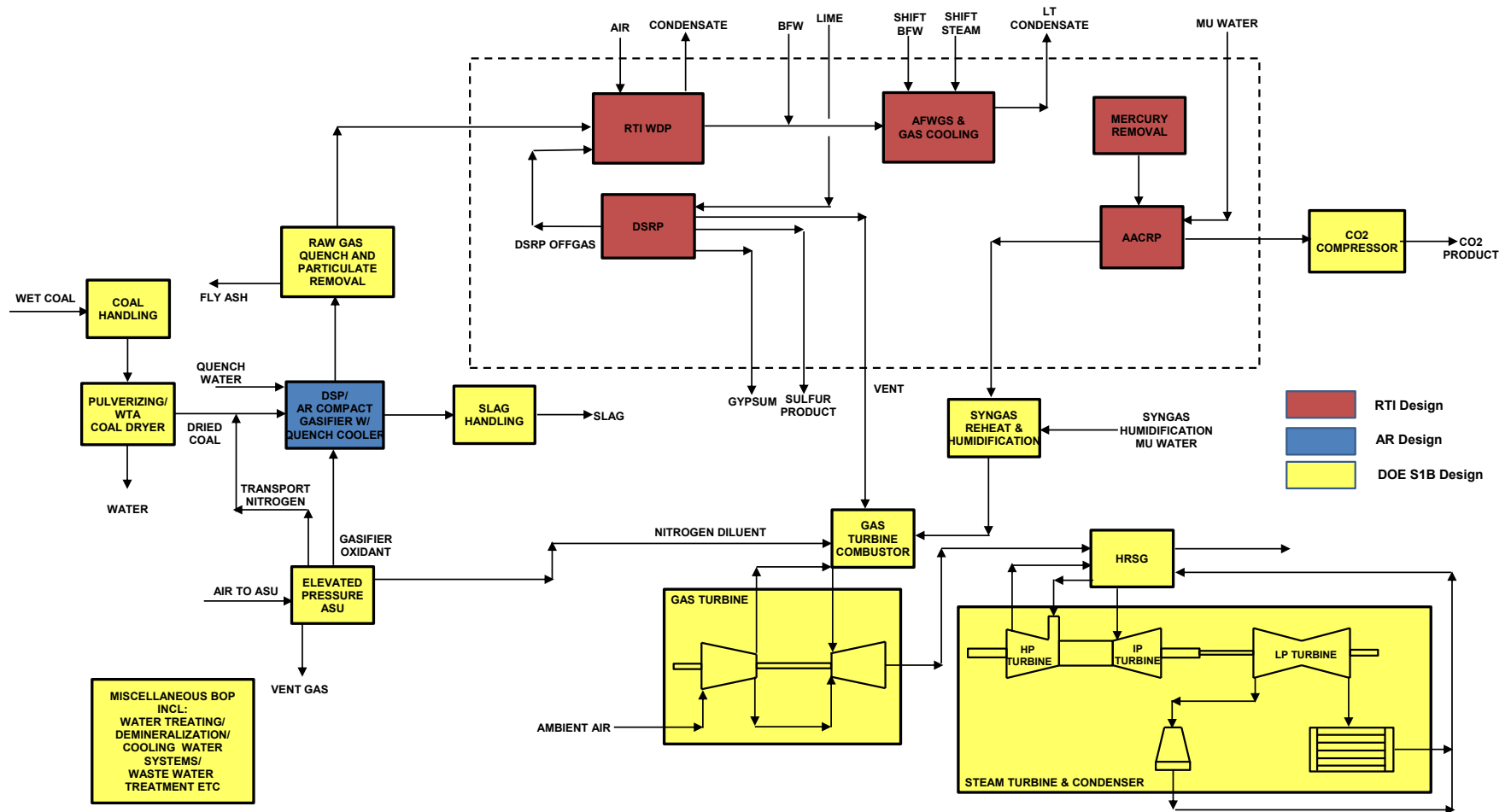
Case 2b is a design that fully integrates the AR advanced gasification with the RTI advanced syngas cleanup and AACRP systems. Due to the different cold gas efficiency of the AR compact gasifier, the IGCC will consume a different amount of coal feed when compared with Case 2a in order to produce nominally, the same amount of hydrogen-rich fuel gas to fill the two advanced GE 7F-turbines and generate 430 MWe at the Montana site. Like Case 2a, the power plant is equipped with a HRSG and ST to generate additional power from waste heat from the flue gas. Due to the different flue gas quantity and IGCC waste heat recovery scheme, the ST output may differ substantially from that of Case 2a.

A combination of RTI's WDP unit and AACRP unit replaces the two-stage Selexol™ unit in Case 2a to remove the sulfur and CO₂ from the syngas. The IGCC plant will remove sulfur via RTI's WDP, which removes hydrogen sulfide (H₂S) and carbonyl sulfide (COS) from the syngas after it leaves the particulate filters, without requiring additional cooling. The treated syngas undergoes sweet shift in RTI's AFWGS before it is cooled and sent to the AACRP unit for CO₂ capture. The AACRP unit will capture CO₂ equivalent to >90% of the raw syngas' carbon content in order to meet the CO₂ emissions specifications.

The Case 2b IGCC plant BFD is shown in Figure 2-2. This figure serves to demarcate the battery limits and highlights the interfaces between AR's and RTI's proprietary systems (colored blue and red respectively) and the rest of the IGCC processes (in yellow) that are Nexant's responsibility for designing. The blue block represents AR's DSP and compact gasifier systems, which replaces the lockhopper feed system and Shell gasifier in the reference Case 2a IGCC plant. The red blocks within the broken-line rectangle represent RTI's advanced syngas cleanup and AACRP processes and comprise the WDP for sulfur removal, DSRP to produce elemental sulfur, AFWGS for high-temperature sweet WGS, low-temperature gas cooling (LTGC), and AACRP for CO₂ removal. The blue and red blocks in Figure 2-2 are the primary focus of AR's and RTI's efforts, and they each provided the system designs for the units within these blocks in Figure 2-2.

The yellow blocks in Figure 2-2 are the Case 2b IGCC systems that are unrelated to the AR advanced gasification and RTI advanced syngas cleanup systems. These blocks are expected to have the same unit operations as those in Case 2a, but with different throughputs due to the different flow quantities. Costs for these systems are scaled based on system capacities from Case 2a using factors given in the QGESS Capital Cost Scaling Methodology document wherever possible.

Figure 2-2
Case 2b: AR Compact Gasifier with RTI WDP IGCC Plant - Simplified BFD



2.2.3 Case 2c: AR Compact Gasifier with Selexol™ AGR IGCC

To complete the evaluation of benefits of the AR and RTI technologies, two other IGCC cases were evaluated, each utilizing either the AR advanced gasification or RTI advanced syngas cleanup on a standalone basis. Case 2c is a design that uses the AR advanced gasification process but paired with the two-stage Selexol™ process to remove both sulfur and CO₂ from the syngas. The coal feed rate is expected to be the same as Case 2b since both utilize the AR gasification process operating under the same conditions, hence same cold gas efficiency, to generate enough syngas for firing in the two GE 7F-turbines.

The Case 2c IGCC plant BFD is shown in Figure 2-3 and the blue block represents AR's DSP and compact gasifier systems. AR provided the design and performance data for these systems. The unchanged yellow blocks are expected to consist of the same unit operations as Case 2a but with different throughputs. Their costs are scaled based on system capacities using factors from the QGESS *Capital Cost Scaling Methodology* document.

2.2.4 Case 2d: Shell Gasifier with RTI WDP IGCC

Case 2d is the design that uses the Shell gasification process paired with RTI advanced syngas cleanup and AACRP processes to remove sulfur and CO₂ from the syngas. The coal feed rate is the same as Case 2a since both utilize Shell gasifiers, which with the same cold gas efficiencies, generate similar syngas quantities for firing in the two GE 7F-turbines to produce 430 MWe of power nominally at the site's elevation.

The red blocks within the broken-line rectangle represent RTI's advanced syngas cleanup process and consist of the unit operations designed by RTI. These include the equipment for WDP, DSRP, AFWGS, LTGC and the AACRP unit. H₂S and COS are removed from the syngas in the WDP unit while primarily CO₂ is removed via the AACRP unit.

The unchanged yellow blocks are expected to consist of the same unit operations as Case 2a but with different throughputs. Like the previous three cases, their costs are scaled based on system capacities using factors from the QGESS *Capital Cost Scaling Methodology* document.

Figure 2-3
Case 2c: AR Compact Gasifier with Selexol™ AGR IGCC - Simplified BFD

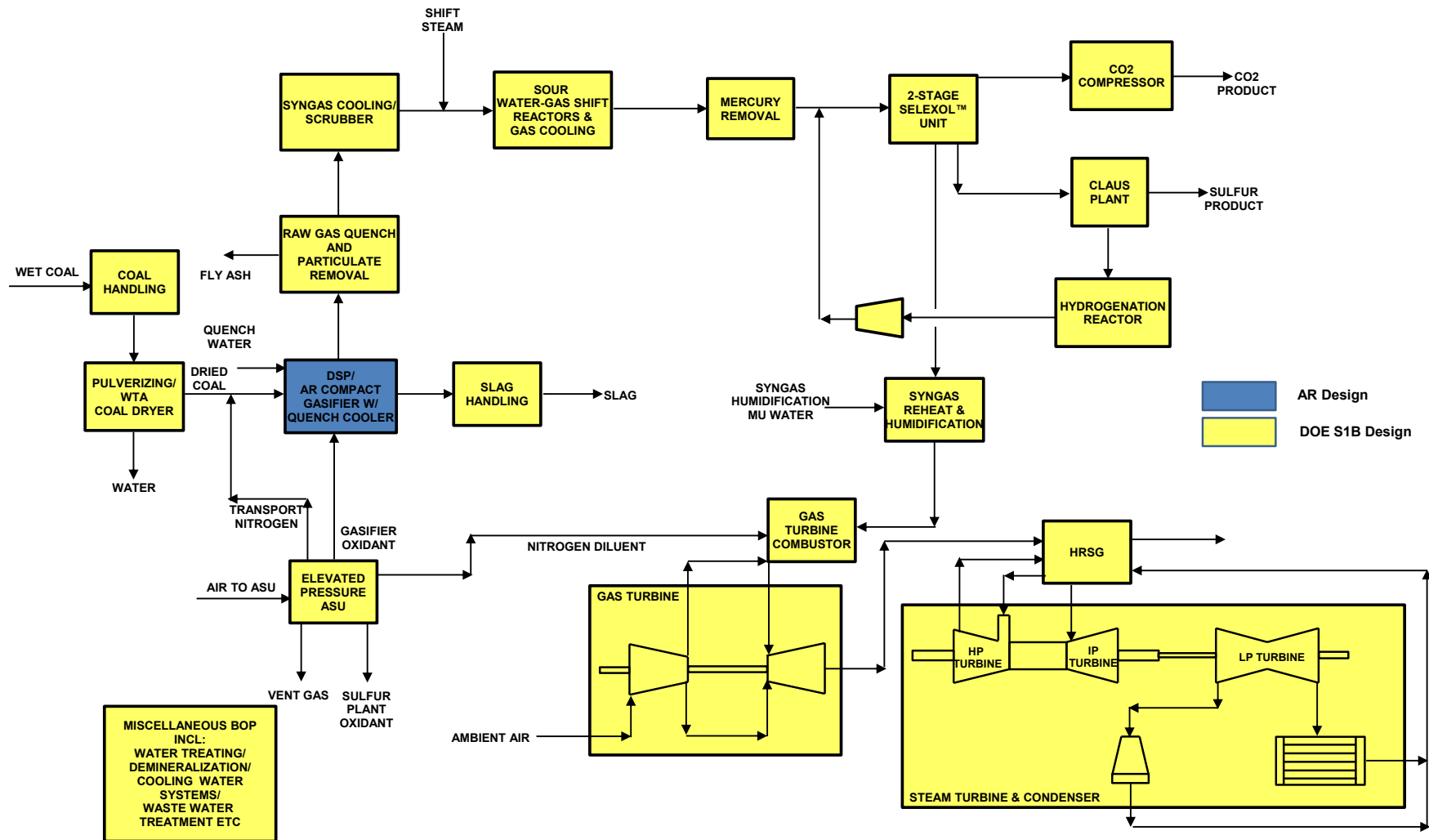
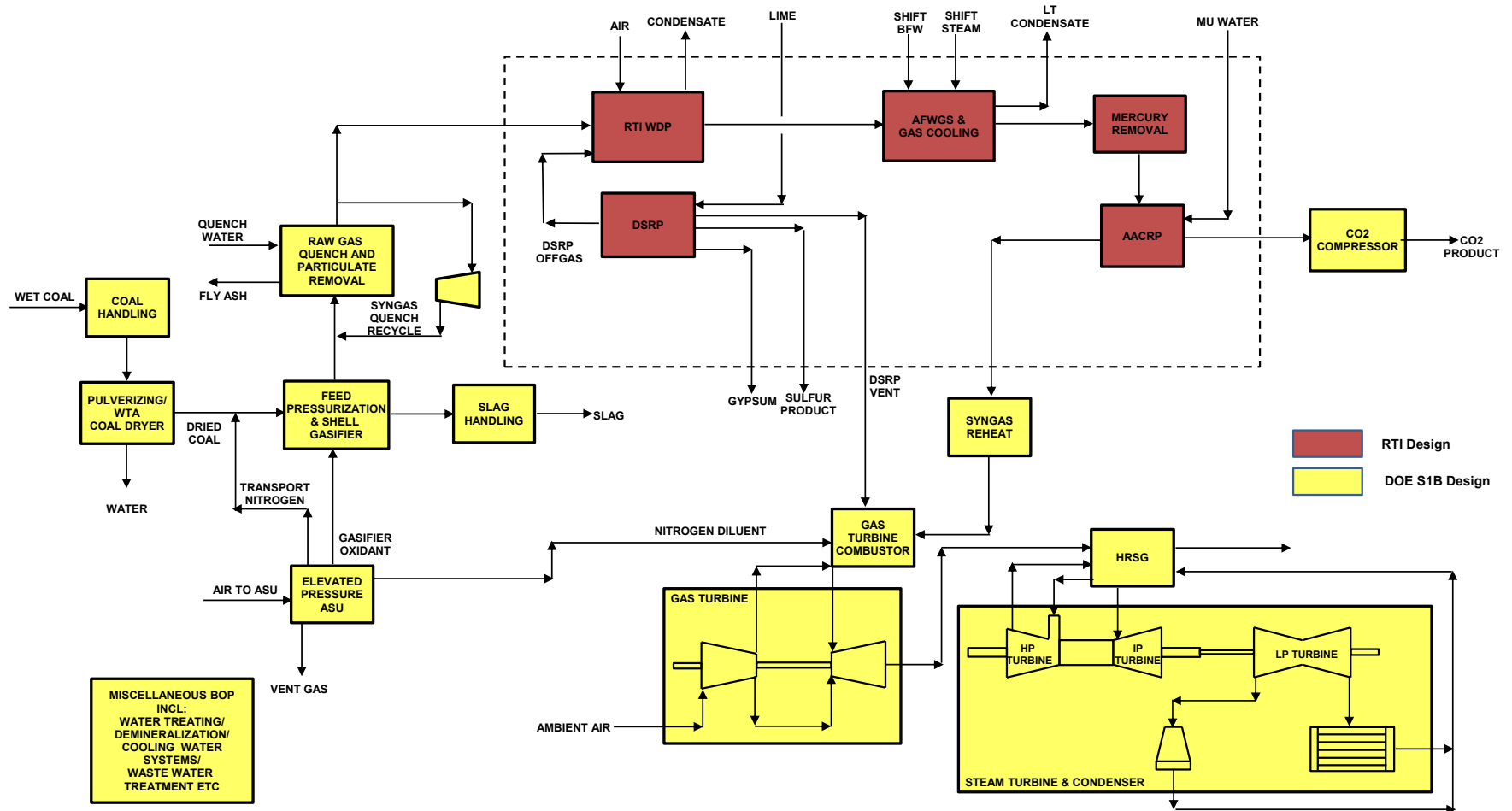


Figure 2-4
Case 2d: Shell Gasifier with RTI WDP IGCC - Simplified BFD



2.3 DESIGN CRITERIA

2.3.1 Heat and Material Balance

Nexant carried out a simulation of all four IGCC cases on ASPEN Plus to obtain the process heat and material balances (HMB). This also facilitated a more detailed estimation of the overall plant utility balance, including heat recovery in the HRSG, power generation from the steam cycle, as well as cooling water load breakdown, all of which helped determine the overall IGCC plant performance with more certainty.

Nexant developed a design of the Case 2a Shell IGCC with Selexol™-based Acid Gas Removal (AGR) using Case S1B data from the NETL Report 1399 as the reference. The resulting stream flows, heat and material balances, and power generation from the gas and steam turbine were benchmarked and cross-checked against the reference Case S1B and the results were within a reasonable range of accuracy.

From the benchmark design, Nexant provided AR and RTI with the stream conditions and flows to the gasification and syngas cleanup units respectively. Using these as inputs into their models, AR and RTI determined the heat and material balances around their respective process systems, overall utilities consumption, and equipment sizes. These outputs were then transmitted to Nexant and plugged into its IGCC simulation to complete the modeling of Cases 2b, 2c, and 2d.

2.3.2 AR Compact Gasifier and Feed System

2.3.2.1 DSP Feed System

AR provided Nexant with the specifications and costs of its proprietary DSP. Nexant had previously completed a TEA (DE-FE0012062) for DOE/NETL on behalf of AR that studied the performance and costs of the overall dry coal feed system. In this study, Nexant integrated the AR DSP with the rest of the feed system equipment, including the atmospheric feed silo, pressurized feed silo, solid filters, nitrogen compressors, etc. to arrive at the overall system cost and performance. For the current study, the performance, utilities consumption and costs for the AR DSP coal feed system were derived from that prior TEA.

2.3.2.2 AR Compact Gasification System

AR provided Nexant with their proprietary compact gasifier's input and output stream flows and conditions, enabling Nexant to integrate the gasifier into its model. AR also provided Nexant the gasifier's utilities consumption and equipment costs. These were used as inputs to help determine the overall IGCC performance and total plant cost (TPC) for Cases 2b and 2c.

2.3.3 RTI WDP System

RTI provided Nexant with the major stream flows and conditions into, within, and out of their proprietary advanced syngas cleanup process and AACRP, enabling Nexant to integrate these processes into its model. For cost estimations of RTI's advanced syngas cleanup and AACRP, RTI provided Nexant with their turnkey costs that were based on actual cost data from the pilot plant and 50 MWe demonstration plant construction. RTI also provided Nexant with a list containing the advanced syngas cleanup and AACRP systems' utilities and consumables rates.

These were used as inputs to establish the overall IGCC performance, TPC and variable operating costs for Cases 2b and 2d.

2.3.4 Other Systems

The costs for the remaining IGCC systems not directly related to the AR advanced gasification and RTI advanced syngas cleanup systems were derived via capacity-factor from the corresponding system costs listed in the reference NETL Report 1399 and escalated to year 2011 costs.

2.4 SITE-RELATED CONDITIONS

The IGCC plants in this study are assumed to be located in Montana, with site-related conditions as shown below:

▪ Location	Montana, US
▪ Elevation, ft above sea level	3,400
▪ Topography	Level
▪ Size, acres	300
▪ Transportation	Rail
▪ Ash/slag disposal	Off Site
▪ Water	Municipal (50%)/Groundwater (50%)
▪ Access	Landlocked, having access by train and highway
▪ CO ₂ disposition	Compressed to 2,215 psia at battery limit before being transported 50 miles for sequestration in a saline formation at a depth of 4,055 ft. (Study scope limited to delivery at IGCC battery limit only)

2.5 METEOROLOGICAL DATA

Maximum design ambient conditions for material balances, thermal efficiencies, system design and equipment sizing are:

▪ Barometric pressure, psia	13.0
▪ Dry bulb temperature (DBT)	42 °F
▪ Wet bulb temperature (WBT)	37 °F
▪ Ambient relative humidity, %	62

2.6 COAL PROPERTIES AND FIRING RATE

Design coal feed to the IGCC power plants is Montana PRB subbituminous coal with characteristics presented in Table 2-2. The as-received coal properties shown in Table 2-2 are from the QGESS Detailed Coal Specifications document. The as-received coal is dried to 6% moisture by the WTA coal drying process (German acronym for “fluidized bed dryer with integrated waste heat recovery”) and fed to the gasifier. The gasifier will gasify enough dried

PRB coal to produce sufficient syngas to fully load two advanced GE 7F-turbines (rated nominally at 215 MWe each) at the Montana site's elevation.

Table 2-2
Montana PRB Coal Specification

Rank	Subbituminous	
Seam	Montana Rosebud PRB	
Source	Western Energy Co.	
Ultimate Analysis, weight%	As-Received	Dried Coal to Gasifier
Carbon	50.07	63.40
Hydrogen	3.38	4.29
Nitrogen	0.71	0.90
Chlorine	0.01	0.01
Sulfur	0.73	0.92
Oxygen	11.14	14.11
Ash	8.19	10.37
Moisture	25.77	6.00
Total	100.0	100.0
Proximate Analysis, weight%	As-Received	Dried Coal to Gasifier
Volatile Matter	30.34	38.42
Fixed Carbon	35.70	45.20
Ash	8.19	10.38
Moisture	25.77	6.00
Total	100.0	100.0
Higher Heating Value (HHV), Btu/lb	8,564	10,825
Sulfur Analysis*, weight%		Dry
Pyritic		0.63
Sulfate		0.01
Organic		0.34
Mercury, ppmW (moisture-free basis)		0.081
Ash Fusion Temperatures at Reducing Conditions, °F		
Initial Deformation	2,238	
Softening	2,254	
Hemispherical	2,270	
Fluid	2,298	

*In accordance with NETL Report 1399, this study assumes that all sulfur in the coal is converted in the gasifier and leaves with the syngas

2.7 CO₂ PRODUCT TREATING AND PURIFICATION DESIGN CRITERIA

For this study, recovered CO₂ is delivered at the battery limit (B/L), with specifications listed in Table 2-3 for saline reservoir sequestration per the QGESS CO₂ Impurities Design Parameters document.

Table 2-3
CO₂ Product Specifications

Compositions:		
CO ₂ ,	vol% (Min)	95
N ₂	vol% (Max)	}4
Ar	vol% (Max)	
O ₂ ,	vol% (Max)	
H ₂	vol% (Max)	
CH ₄ ,	vol% (Max)	
H ₂ O,	ppmv (Max)	300
H ₂ S	vol% (Max)	0.01
NH ₃	ppmv (Max)	50
SO ₂ ,	ppmv (Max)	100
NO _x ,	ppmv (Max)	100
CO,	ppmv (Max)	35
B/L pressure, psia		2,215
B/L Temperature, °F		95

2.8 POWER GENERATION & AUXILIARY LOADS

The NETL Report 1399 provided a breakdown of the Case S1B (Shell gasification-based IGCC with CO₂ Capture) power generation by gas and steam turbine power generation. It also provided the auxiliary loads for Case S1B, broken down into its major systems. For the power generation portion in each of the four cases in the current study, Nexant provided a revision using its model, which is based on Case S1B IGCC plant's power generation design. Auxiliary loads were estimated, wherever applicable, by pro-rating from the S1B case using relevant scaling parameters obtained from the model's heat and material balance.

Table 2-4 shows the power production and auxiliary load breakdown of the original DOE/NETL Case S1B from the NETL Report 1399, which Case 2a of this study was modeled upon and benchmarked against. For reference purposes, the scaling parameters are also shown in the table.

The AR advanced gasification system and RTI advanced syngas cleanup and AACRP systems have different auxiliary loads that are not scalable with some of the loads specified in the reference Case S1B. AR and RTI have provided the auxiliary loads for their systems and these were used directly as inputs to the auxiliary load calculation.

Table 2-4
DOE S1B IGCC Power Production and Auxiliary Load Summary

Power Generation	Electrical Load, kWe	Scaling Capacity
Gas Turbine Power	430,900	Model Output
Steam Turbine Power	232,500	Model Output
TOTAL POWER, kWe	663,400	
Auxiliary Load Description	Electrical Load, kWe	Scaling Capacity
Coal Handling	510	As Received Coal
Coal Milling	2,730	As Received Coal
Slag Handling	580	Slag Flow
WTA Coal Dryer Compressor	9,370	Moisture Removed
WTA Coal Dryer Auxiliaries	620	Moisture Removed
Air Separation Unit Auxiliaries	1,000	Oxygen Production
Air Separation Unit Main Air Compressor	63,550	Oxygen Production
Oxygen Compressor	8,830	Oxygen Production
Nitrogen Compressors	33,340	N ₂ Diluent + Conveying N ₂
CO ₂ Compressor	31,560	CO ₂ Product Flow
Boiler Feedwater Pumps	3,260	Boiler Feed Water (BFW) Flow
Condensate Pump	230	Condensate Flow
Quench Water Pump	760	Quench Water Flow
Syngas Recycle Compressor	820	Recycle Syngas Flow
Circulating Water Pump	2,730	Circulating Water Flow
Ground Water Pumps	310	Circulating Water Flow
Cooling Tower Fans	1,780	Cooling Water Flow
Air Cooled Condenser Fans	2,960	Condenser Duty
Scrubber Pumps	20	Syngas Flow to Scrubber
Acid Gas Removal	18,400	CO ₂ + H ₂ S Flow
Gas Turbine Auxiliaries	1,000	Gas Turbine Power Output
Steam Turbine Auxiliaries	100	Steam Turbine Power Output
Claus Plant/TGTU Auxiliaries	250	H ₂ S Flow
Claus Plant TG Recycle Compressor	1,530	H ₂ S Flow
Miscellaneous Balance of Plant	3,000	As Received Coal
Transformer Losses	2,550	Total Power Output
TOTAL AUXILIARIES, kWe	191,790	
NET POWER, kWe	471,610	

All comparison cases completed within this study have the same basis. Because of this approach, the relative differences between the cases enables comparison of performance and costs and identifies potential trends for the different technologies being evaluated. Unfortunately, the general audience may be more familiar with results from similar studies, but conducted at the more typical reference elevation of sea level. Because a full evaluation of the performance of any selected IGCC case from this study at a sea level elevation was outside of this project's scope,

the following preliminary analysis was completed to provide an estimated of the performance at sea level. Based on the available information on GE F7-turbines, the estimated output at sea level should be 230 MWe. Similarly, the auxiliary loads associated with any compressor or pump, which compresses or pumps fluids from ambient conditions to system operating pressure, will also change for a plant at sea level. Because the difference in barometric pressure between the two elevations and typical system operating pressure is less than 0.3 %, the net auxiliary load can for a first approximation be assumed to equal to that evaluate at the design elevation for the study. Similarly, the estimated costs have been assumed to be equivalent to the current study. These assumptions have been used to generate the preliminary comparison data provided in Table 2-5 for reference Case 2a from this study.

Table 2-5
Performance Comparison for Reference Plant at Two Different Elevations

Performance Metric	Reference Case	
	Elevation: 3,400 ft	Elevation: Sea level
Power Production, kWe		
Gas Turbine	429,973	464,000
Steam turbine	224,080	224,080
Auxiliary Power Consumption ¹ , kWe	193,448	193,448
Net Power Output, kWe	460,605	494,632
Total Overnight Cost (2011\$MM)	2,487	2,487
COE, excl CO ₂ TS&M ^{2,3} , mills/kwh	145.3	135.4
COE, incl CO ₂ TS&M ^{2,3} , mills/kwh	166.2	154.8
Cost of CO ₂ Avoided excl CO ₂ TS&M ^{2,3} , \$/ton CO ₂	79.8	67.8
Cost of CO ₂ Avoided incl CO ₂ TS&M ^{2,3} , \$/ton CO ₂	105.1	91.3

¹ Auxiliary loads at sea level have been assumed to be equal to results at 3,400 ft of elevation. The loads will change, but fully evaluating these changes was beyond the scope of this project.

² The costs from the reference case were escalated to 2011\$ based on DOE escalation for GE IGCC case with carbon capture in DOE Report, “Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases”.

³ Reference case cost of electricity (COE) and CO₂ avoided costs at sea level were estimated by RTI.

2.9 RAW WATER SUPPLY

Raw water makeup was assumed to be provided 50% by a publicly owned treatment works and 50% from groundwater.

2.10 ENVIRONMENTAL DESIGN TARGETS

The IGCC emission targets were established in the Electric Power Research Institute’s (EPRI) design basis for their CoalFleet for Tomorrow Initiative, documented in “CoalFleet User Design Basis Specifications for Coal-Based IGCC Power Plants, EPRI, Palo Alto, CA, 2009.” The design targets were established specifically for bituminous coal but apply to subbituminous coals as well. The emissions requirements and limits for the reference IGCC power plant, as specified in NETL Report 1399, are listed in Table 2-6 below:

Table 2-6
IGCC Environmental Targets

Pollutant	Environmental Target	NSPS Limit
NO _x	15 ppmv (dry) @ 15% O ₂	1.0 lb/MWh
SO ₂	0.0128 lb/MMBtu	1.4 lb/MWh
Particulate Matter (PM)	0.0071 lb/MMBtu	0.015 lb/MMBtu
Hg	>90% capture	20 x 10 ⁻⁶ lb/MWh

Total air pollutants in all vents must meet the above specifications even if atmospheric venting is minimal.

2.11 OTHER SITE SPECIFIC REQUIREMENTS

Although the following design parameters are considered site-specific, and are not quantified for this study, allowances for normal conditions and construction are included in the cost estimates.

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Fire protection
- Local code height requirements
- Noise-regulations/impact on site and surrounding area

2.12 CAPITAL COST ESTIMATION METHODOLOGY

2.12.1 General

For IGCC plants with CO₂ capture, the NETL Report 1399 provided a code of accounts grouped into 14 major systems. Each of these major systems is broken down further into different subsystems. This type of code of accounts structure has the advantage of grouping all reasonably allocable components of a system or process into a specific system account.

For the four IGCC cases evaluated in this study, capital cost scaling following the guidelines and parameters that are described in the NETL Capital Cost Scaling Methodology document was used to perform the cost estimation for systems that are not related to the AR advanced gasification and RTI advanced syngas cleanup and AACRP systems. In general, this cost estimation methodology involves determining the scaling parameters, exponents and coefficients from the Capital Cost Scaling Methodology, as well as the reference cost and baseline capacity from the NETL Report 1399. Once these have been established, the capital cost can be estimated

based on the revised capacity from the HMB developed by Nexant's ASPEN models of the IGCC cases.

Nexant performed a bottoms-up, major equipment (ME) factored cost estimation for the AR advanced gasification and RTI advanced syngas cleanup and AACRP systems. These are described in greater detail in Section 2.12.2.

2.12.2 Advanced Technology System Capital Cost Estimate Criteria

2.12.2.1 *AR-Licensed Systems Cost Estimates*

Costs for equipment in the AR advanced gasification system that are proprietary to AR, such as its DSP and compact gasifier, were provided by AR. Nexant used and reported these costs on an as-provided basis. The level of detail provided in these cost estimates was determined by AR itself.

2.12.2.2 *Nexant-Designed System Cost Estimates*

For auxiliary equipment within the AR advanced gasification section that is not included in the technology licensor's package, Nexant performed a bottom-up cost estimate based on equipment sizing. Based on the equipment sizes defined by process HMB, cost for each piece of major equipment was estimated based on either vendor quotes or using commercial estimation software (ASPEN In-Plant Cost Estimator). Installation labor hours for the equipment were factored from historical data by equipment type.

As defined in the DOE Report 1399, an average labor wage at \$39.7/hr, with an all-in labor cost of \$51.6/hr (including wages plus 30% burden to cover fringe benefits, payroll based taxes, and insurance premiums) was assumed for calculating the 2011 installation labor costs. No over-time or other premiums were added. The average labor productivity for the site was assumed to be 105% of the US Gulf coast productivity.

Bulk material and installation costs were factored from the major equipment cost (MEC). Bulk materials, which include instrumentation, piping, structure steel, insulation, electrical, painting, concrete & site preparation works, and are needed to complete the major equipment installations were factored from MEC based on historical data for similar services. Installation labor for each bulk commodity was factored from historical data by type. Sum total of MEC plus bulk material costs plus installation labor costs forms the total direct cost (TDC).

Construction indirect costs were then factored from total direct labor costs based on historical data, and added to the system TDC to give the total field cost (TFC) for the system. Construction indirect cost covers the cost for setup, maintenance and removal of temporary facilities, warehousing, surveying and security services, maintenance of construction tools and equipment, consumables and utilities purchases, and field office payrolls. It should be noted that the term TFC is the equivalent of the Bare Erected Cost (BEC) used in the DOE Report 1399.

2.12.3 RTI Advanced syngas Cleanup and AACRP Systems Cost Estimates

For WDP, RTI's process model incorporates actual data from the pilot plant testing at Eastman Chemical and the 50 MWe demonstration plant at Tampa Electric Company. For AACRP, RTI

developed a detailed process model using ProMax with TSWEET kinetics for the CO₂-amine reactions. This process model was reviewed by a recognized expert in the field of acid gas removal technologies. RTI has diligently worked to validate these models against actual pilot and demonstration plant data for RTI's proprietary technology and any publicly available DOE reports for the more commercial technologies.

Using these models, RTI developed HMBs and sized equipment lists. Using process conditions and experience gained from the pilot plant and demonstration plant operation, materials of construction were identified. With the sized equipment list and materials of construction, RTI developed equipment cost estimates from actual vendor quotes or scaled equipment costs from the 50 MWe demonstration unit for a majority of the equipment. For some of the more generic and non-proprietary equipment, the equipment costs were developed using Aspen Plus ICARUS Economic Analyzer. RTI then used the same estimating factors used in NETL Report 1399 to estimate BEC for WDP, AFWGS, and AACRP. This BEC information was used by Nexant directly in its calculation of the overall IGCC capital expenditure (CAPEX).

In addition to using the actual pilot plant and demonstration plant data for developing equipment costs, RTI used this available data to develop utility and variable cost especially for sorbent, catalysts, and solvent. Estimates of the utilities and variable costs were also provided to Nexant. Nexant used these to complete additional performance analysis and calculate the IGCC plant's overall variable operating costs.

2.12.4 Balance of Plant Capital Cost Estimate Criteria

The capital cost estimates for the rest of the IGCC systems that are unrelated to coal gasification, sulfur and CO₂ removal, were developed based on the Case S1B Shell IGCC plant with CO₂ capture case in the NETL Report 1399. The costs were adjusted for differences in unit or plant capacity according to NETL's Guidelines as described in the NETL Capital Cost Scaling Methodology document.

Table 2-7 shows the code of accounts for the IGCC plant. These systems are further broken down to include the various subsystems. The scaling parameters for these Balance of Plant (BOP) subsystems, as laid out by the NETL Capital Cost Scaling Methodology document, are also shown in this table.

Table 2-7
Code of Accounts for Report IGCC Plant

Acct No.	Item/Description	Scaling Parameter
1	COAL & SORBENT HANDLING	
1.1	Coal Receive & Unload	Coal Feed Rate
1.2	Coal Stackout & Reclaim	Coal Feed Rate
1.3	Coal Conveyors & Yard Crush	Coal Feed Rate
1.4	Other Coal Handling	Coal Feed Rate
1.9	Coal & Sorbent Handling Foundations	Coal Feed Rate
2	COAL & SORBENT PREP & FEED	
2.1	Coal Crushing & Drying	Coal Feed Rate
2.2	Prepared Coal Storage & Feed	Coal Feed Rate
2.3	Dry Coal Injection System	Calculated
2.4	Misc Coal Prep & Feed	Coal Feed Rate
2.9	Coal & Sorbent Feed Foundation	Coal Feed Rate
3	FEEDWATER & MISC. BOP SYSTEMS	
3.1	Feedwater System	BFW (HP only)
3.2	Water Makeup & Pretreating	Raw Water Makeup
3.3	Other Feedwater Subsystems	BFW (HP only)
3.4	Service Water Systems	Raw Water Makeup
3.5	Other Boiler Plant Systems	Raw Water Makeup
3.6	FO Supply Sys and Nat Gas	Coal Feed Rate
3.7	Waste Treatment Equipment	Raw Water Makeup
3.8	Misc Power Plant Equipment	Coal Feed Rate
4	GASIFIER & ACCESSORIES	
4.1	Gasifier, Syngas Cooler & Auxiliaries	Syngas Throughput
4.3	ASU/Oxidant Compression	O ₂ Production
4.4	Scrubber & Low Temperature Cooling	Syngas Flow
4.6	Other Gasification Equipment	Syngas Flow
4.9	Gasification Foundations	Syngas Flow
5A	GAS CLEANUP & PIPING	
5A.1	Double Stage Selexol™	Gas Flow to AGR
5A.2	Elemental Sulfur Plant	Sulfur Production
5A.3	Mercury Removal	Hg Bed Carbon Fill
5A.4	Shift Reactors/COS Hydrolysis	WGS/COS Catalyst
5A.5	Blowback Gas Systems	Candle Filter Flow
5A.6	Fuel Gas Piping	Fuel Gas Flow
5A.9	HGCU Foundations	Sulfur Production
5B	CO ₂ REMOVAL & COMPRESSION	
5B.2	CO ₂ Compression & Drying	CO ₂ Flow
6	COMBUSTION TURBINE/ACCESSORIES	
6.1	Combustion Turbine Generator	Fuel Gas Flow
6.2	Combustion Turbine Foundations	Fuel Gas Flow
7	HRSG, DUCTING & STACK	
7.1	Heat Recovery Steam Generator	HRSG Duty
7.3	Ductwork	Vol Flow to Stack
7.4	Stack	Vol Flow to Stack
7.9	HRSG, Duct & Stack Foundations	Vol Flow to Stack

Acct No.	Item/Description	Scaling Parameter
8	STEAM TURBINE GENERATOR	
8.1	Steam TG & Accessories	Turbine Capacity
8.2	Turbine Plant Auxiliaries	Turbine Capacity
8.3a	Condenser & Auxiliaries	Condenser Duty
8.3b	Air Cooled Condenser	Condenser Duty
8.4	Steam Piping	BFW (HP Only)
8.9	TG Foundations	Turbine Capacity
9	COOLING WATER SYSTEM	
9.1	Cooling Towers	Cooling Tower Duty
9.2	Circulating Water Pumps	Circ H ₂ O Flow Rate
9.3	Circ. Water System Auxiliaries	Circ H ₂ O Flow Rate
9.4	Circ Water Piping	Circ H ₂ O Flow Rate
9.5	Makeup Water System	Raw Water Makeup
9.6	Component Cooling Water System	Circ H ₂ O Flow Rate
9.9	Circ. Water System Foundations	Circ H ₂ O Flow Rate
10	ASH/SPENT SORBENT HANDLING SYS	
10.1	Slag Dewatering & Cooling	Slag Production
10.6	Ash Storage Silos	Slag Production
10.7	Ash Transport & Feed Equipment	Slag Production
10.8	Misc. Ash Handling System	Slag Production
10.9	Ash/Spent Sorbent Foundation	Slag Production
11	ACCESSORY ELECTRIC PLANT	
11.1	Generator Equipment	Turbine Capacity
11.2	Station Service Equipment	Auxiliary Load
11.3	Switchgear & Motor Control	Auxiliary Load
11.4	Conduit & Cable Tray	Auxiliary Load
11.5	Wire & Cable	Auxiliary Load
11.6	Protective Equipment	Auxiliary Load
11.7	Standby Equipment	Total Gross Output
11.8	Main Power Transformers	Total Gross Output
11.9	Electrical Foundations	Total Gross Output
12	INSTRUMENTATION & CONTROL	
12.4	Other Major Component Control	Auxiliary Load
12.6	Control Boards, Panels & Racks	Auxiliary Load
12.7	Computer & Accessories	Auxiliary Load
12.8	Instrument Wiring & Tubing	Auxiliary Load
12.9	Other I & C Equipment	Auxiliary Load
13	IMPROVEMENT TO SITE	
13.1	Site Preparation	Accounts 1-12
13.2	Site Improvements	Accounts 1-12
13.3	Site Facilities	Accounts 1-12
14	BUILDING & STRUCTURES	
14.1	Combustion Turbine Area	Gas Turbine Power
14.2	Steam Turbine Building	Accounts 1-12
14.3	Administration Building	Accounts 1-12
14.4	Circulation Water Pumphouse	Circ H ₂ O Flow Rate
14.5	Water Treatment Buildings	Raw Water Makeup
14.6	Machine Shop	Accounts 1-12
14.7	Warehouse	Accounts 1-12
14.8	Other Buildings & Structures	Accounts 1-12
14.9	Waste Treating Building & Structures	Raw Water Makeup

As Table 2-7 is based on DOE/NETL Report 1399, it does not necessarily have account and/or subaccount numbers for the advanced technologies being evaluated in this study. To support optimal comparison of these advanced technologies with the existing commercial technologies, the advanced technologies used the same account and/or subaccount numbers as the existing commercial technologies that they are most analogous to. For technologies without a defined account number, one was created. Table 2-8 provides a list of the advanced technologies evaluated in this study and their associated account numbers.

Table 2-8
Code of Accounts for Advanced Technologies being Evaluated

Account Number	Title	Advanced Technology
2.3	Dry Coal Injection System	AR DSP
4.1	Gasifier, Syngas Cooler & Auxiliaries	AR Compact Gasifier
5A.1	RTI WDP	RTI WDP
5A.2	RTI DSRP	RTI DSRP
5A.4	Shift Reactors	AFWGS
5B.1	AACRP	AACRP

2.12.5 Home Office, Engineering Fees and Project/Process Contingencies

Engineering and construction management fees and home office cost, project and process contingencies were factored from the each subsystem's TFC. These were then added to the TFC to come up with the TPC of the system. Factors from Case S1B in the NETL Report 1399 were used.

2.12.6 Owner's Cost

Owner's cost was then added to TPC to come up with the total overnight cost (TOC) for the system. Owner's costs as defined in the NETL Report 1399 include the following:

- Preproduction costs –
 - o 6 months of all labor cost
 - o 1 month of maintenance materials
 - o 1 month of non-fuel consumables
 - o 1 month of waste disposal
 - o 25% of 1 month fuel cost at 100% CF
 - o 2% TPC
- Inventory capital -
 - o 60 day supply of fuel and consumable at 100% CF
 - o 0.5% TPC
- Initial cost for catalyst and chemicals per design
- Land cost = \$900,000 (300 acres x \$3,000/acre)
- Other owner's costs at 15% TPC

- Financing costs at 2.7% TPC

2.13 OPERATION & MAINTENANCE COSTS

The operation and maintenance (O&M) costs pertain to those charges associated with operating and maintaining power plants over their expected life. These costs include:

- Operating labor
- Maintenance – material and labor
- Administrative and support labor
- Consumables
- Fuel
- Waste disposal

There are two components of O&M costs; fixed O&M, which is independent of power generation, and variable O&M, which is proportional to power generation. Variable O&M costs were estimated based on an 80% CF.

2.13.1 Fixed Costs

Operating labor cost was determined based on the number of operators required to work in the plant. Other assumptions used in calculating the total fixed cost include:

• 2011 base hourly labor rate, \$/hr	\$39.7
• Length of work-week, hr	50
• Labor burden, %	30
• Administrative/support labor, % O&M labor	25
• Maintenance material + labor, % TPC	2.8
• Maintenance labor only, % maintenance material + labor	35
• Property taxes and insurances, % TPC	2

2.13.2 Variable Costs

The cost of consumables, including fuel, was determined based on the individual rates of consumption, the unit cost of each specific consumable commodity, and the plant annual operating hours. Waste quantities and disposal costs were evaluated similarly to the consumables.

The unit costs for major consumables and waste disposal was selected from NETL Report 1399, QGESS Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases and from the QGESS Fuel Prices for Selected Feedstocks in NETL Studies documents.

The 2011 coal price as delivered to the Montana IGCC plant was \$19.63/ton, per the QGESS Fuel Prices for Selected Feedstocks in NETL Studies document.

2.13.3 CO₂ Transport and Storage Costs

As specified in NETL Report 1399, CO₂ Transport, Storage and Monitoring (T&SM) costs used for the Montana IGCC plant location is \$22/tonne. The Cost of Electricity (COE) is reported both with and without the cost of CO₂ TS&M.

2.14 FINANCIAL MODELING BASIS

2.14.1 Cost of Electricity

The primary metric used to evaluate overall financial performance is the COE for the IGCC plant. All costs were expressed in the “first-year-of-construction” year dollars, and the resulting COE was also expressed in “first-year-of-construction” year dollars.

The same financial modeling methodology was used for this study as per the NETL Report 1399 and guidelines in the QGESS Cost Estimation Methodology for NETL Assessments of Power Plant Performance document. This is a simplified method that is a function of the plant TPC, capital charge factor (CCF), fixed and variable operating costs, CF and net power generation, as shown in the equation below:

$$COE = \frac{\text{first year capital charge} + \text{first year fixed operating costs} + \text{first year variable operating costs}}{\text{annual net megawatt hours of power generated}}$$
$$COE = \frac{(CCF)(TOC) + OC_{FIX} + (CF)(OC_{VAR})}{(CF)(MWH)}$$

The CCF used in evaluating the COE was pre-calculated using the NETL Power Systems Financial Model (PSFM). This factor is valid for global economic assumptions used for a pre-determined finance structure and capital expenditure period. For the IGCC with CO₂ capture cases, the financial performance evaluations are in accordance with the high-risk, Investor Owned Utility (IOU) finance structure with a 5 year capital expenditure period. The resulting CCF is 0.1243.

2.14.2 Cost of CO₂ Avoided

Per the reporting requirements of the TEA, the cost of CO₂ avoided shall be reported, if a reference non-capture plant is available. Since the scope of work did not specify the modeling of an analogous case without capture, a reference baseline supercritical (SC) pulverized coal (PC) boiler plant firing the same coal at the same design conditions is used instead, as recommended by the NETL Report 1399. The equation for the CO₂ avoided cost is:

$$\text{Avoided Cost} = \frac{\{COE_{\text{with removal}} - COE_{\text{reference}}\} \$ / MWh}{\{CO_2 \text{ Emissions}_{\text{reference}} - CO_2 \text{ Emissions}_{\text{with removal}}\} \text{ tons} / MWh}$$

The baseline SC PC plant firing PRB coal is a 550 MWe plant with results presented in Table 2-9.

Table 2-9
Baseline SC PC Results for CO₂ Avoided Cost Calculation

	PRB Coal
Net Output, MW	550
*2011 COE, mills/kWh	79.42
Emissions, lb/net-MWh	1,892

*The 2011 COE for 550 MW PRB coal-fired SC PC plant was escalated to 2011 by applying an escalation factor to the 2007 COE of 57.80. This factor is obtained by dividing the 2011 550 MW Illinois #6 coal fired SC PC COE by the corresponding 2007 COE (80.95/58.91)*57.8=79.42

2.14.3 CO₂ Sales Price

Sensitivity analysis was used to establish the impact of CO₂ sales on IGCC COE. The CO₂ sales price at the IGCC plant gate was varied between \$0/tonne (baseline case assuming no value to the product CO₂) and \$60/tonne.

The formula used to calculate the revised COE after taking into account CO₂ sales was

$$COE = \text{Baseline COE} - \frac{(\text{CO}_2 \text{ Sales Price}) \times \text{annual tonnes of CO}_2 \text{ product}}{\text{annual net megawatt hours of power generated}}$$

2.14.4 Cost of CO₂ Emissions

Sensitivity analysis on cost of CO₂ emissions was also performed using the CO₂ emissions cost as a variable. The emissions cost ranged between \$0/tonne (baseline case assuming no CO₂ emissions cost) and \$60/tonne.

The formula used to calculate the revised COE after taking into cost of CO₂ emissions was

$$COE = \text{Baseline COE} + \frac{(\text{Cost of CO}_2 \text{ Emissions}) \times \text{annual tonnes of CO}_2 \text{ emitted}}{\text{annual net megawatt hours of power generated}}$$

3.1 PROCESS OVERVIEW

The reference Case 2a IGCC power plant, which is based on the NETL Report 1399 Case S1B, is a Montana PRB coal-fired IGCC plant designed to generate enough hydrogen-rich fuel gas to fill two advanced GE 7F-turbines rated nominally at 215 MWe each for a total of 430 MWe at the Montana site's elevation. The power plant is equipped with a HRSG and ST to maximize power recovery. It is designed to capture CO₂ equivalent to 90% of the raw syngas' carbon content using the two-stage Selexol™ process. The nominal net IGCC power export capacity after accounting for the auxiliary loads which include CO₂ capture and compression is 461 MWe.

In order to achieve the 90% CO₂ removal target and maintain the same syngas heat content (Btu/SCF) to the gas turbine (GT), the raw syngas must be converted to hydrogen-rich syngas by the WGS reaction. Steam for the WGS reaction is provided partly by the steam generated from quench cooling of the syngas and partly by the water content in scrubber overhead gas that is also feeding to the unit. Additional steam is injected to the syngas stream to push the equilibrium towards a high conversion of CO. The WGS catalyst also hydrolyzes the COS to H₂S for capture in the AGR process. The recovered H₂S is converted into elemental sulfur in the Claus plant.

The IGCC plant is assumed to operate as a base-loaded unit with annual on-stream CF of 80% or 7,000 hrs/year at full capacity. This capacity factor was selected as it represents the maximum availability demonstrated by commercial IGCC plants.

3.2 IGCC COMMON PROCESS AREAS

As shown in Figure 2-1, the Case 2a IGCC power plant consists of the following major process and/or utility blocks. Some of these blocks, or process areas, are common to the Case 2b, 2c, and 2d plant configurations. These common process areas are in bold and italicized.

- Coal Sizing Handling
- Coal Prep, Drying
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- Dry Coal Feed & Shell SCGP Gasifier System
- Syngas Cooling (Quench, Scrubbing, Steam Generation)
- Gas Cleaning (Filters, WGS & AGR)
- Mercury Removal
- CO₂ Compression and Purification Facilities
- Sulfur Plant
- Combustion Turbine Power Generation (CTG)
- HRSG, Ducting and Stack
- Steam Turbine Power Generation (STG)
- Cooling Water Systems

-
- BFW/Condensate System
 - Slag Recovery and Handling
 - Accessory Electric Plant
 - Instrumentation and Control

The common areas are presented in brief here for general background information, and to avoid unnecessary repetition in the other cases. Detailed descriptions of these process areas can be found in Section 3.1 of the NETL Report 1399. Where there is case specific performance information, these features are presented in the relevant case sections.

3.2.1 Coal Sizing and Handling

The PRB coal is delivered to the site by 100-ton rail cars. It is unloaded into two receiving hoppers and fed to the vibratory feeder. It is then transferred through intermediate hoppers and silos to the coal crusher where it is reduced to 1-1/4" x 0 size.

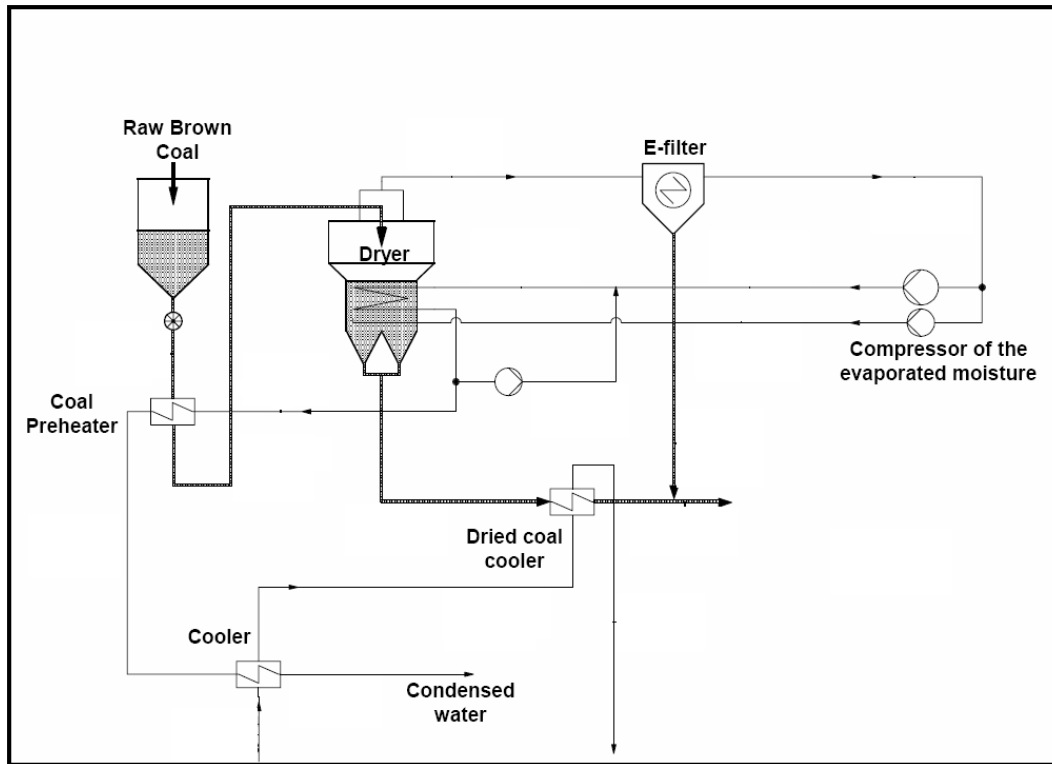
3.2.2 Coal Preparation and Drying

A paper presented by Shell in the Gasification Technology Conference was cited in the Reference S1B case. The paper described the drying of subbituminous coal to 6% moisture before feeding it to the Shell entrained-flow gasifier. This moisture content is considered compatible with the storage, transport, and feed injection requirements for the Shell entrained flow gasifier.

The coal drying process selected in the NETL Report 1399 is the fine grain WTA process. It was chosen for its ability to recover the coal moisture for use in a closed-loop drying process instead of discharging the moisture to the atmosphere and that syngas is not required to provide heat for coal drying. A process schematic is shown in Figure 3-1.

AR has indicated that its DSP and compact gasifier are compatible with the PRB coal that has been dried to 6% moisture by a WTA dryer. Hence, for Cases 2b and 2c, the WTA coal drying system is still the process of choice.

Figure 3-1
WTA Coal Drying Process Schematic



3.2.3 ASU

The reference Case S1B utilizes an “elevated pressure” ASU in which the main air compressor discharge pressure is 190 psia. No air supply integration with the GT compressor is used. In addition to providing 95% oxygen to the gasifiers and the Claus plant, the ASU also provides diluent nitrogen to the GT combustor to increase GT power output, maintaining optimum firing temperatures and minimize the formation of NO_x .

The battery limit conditions for the ASU products are summarized in Table 3-1.

Table 3-1
ASU Product Conditions

ASU Product	Pressure, psia	Temperature, °F
95% O_2	125	90
Diluent N_2	384	385
Transport N_2	815	387
ASU Vent	16.4	64

3.2.4 Mercury Removal

Mercury is removed from the syngas at elevated pressure and prior to combustion because syngas volumes are much smaller than flue gas volumes. Mercury removal is achieved via an activated carbon bed adsorption system as cited in the NETL Report 1399. Data on the performance of carbon bed systems was obtained from the Eastman Chemical Company, which uses carbon beds at its syngas facility in Kingsport, Tennessee.

3.2.5 CO₂ Compression and Dehydration

A raw CO₂ product with greater than 99% purity leaves the Selexol™ AGR plant at the conditions of 150 psia and 60°F. It is compressed to supercritical condition of 2,215 psia using a multi-stage, intercooled compressor. The CO₂ stream is dehydrated to a dew point of -40 °F at the appropriate inter-stage pressure using a thermal swing adsorptive dryer.

3.2.6 Slag and Ash Handling

Slag material drains from the gasifier into a water bath in the bottom of the gasifier vessel. The slag-water slurry is transferred to a slag crusher where the slag is crushed into pea size fragments. The slurry containing 5 to 10% solids is then transferred to a dewatering bin through a lockhopper for dewatering. The water is clarified and reused as makeup to the water scrubber. The dried slag is stored for disposal.

3.2.7 Combustion Turbine Power Generation

The GT generator selected is an advanced F class turbine. Nitrogen from the ASU is used for dilution to limit NO_x formation and to adjust the syngas Lower Heating Value (LHV) to 115-132 Btu/SCF. Inlet air is compressed to a pressure ratio of 16:1 for the GT combustion process. Hot combustion products are expanded in a three stage turbine expander with a last stage exhaust temperature of around 1,050°F. The nominal gross GT output per turbine is 215 MWe at the Montana site location.

3.2.8 Steam Turbine and HRSG

The 1,050°F GT exhaust is cooled in the HRSG by generating High Pressure (HP), Intermediate Pressure (IP), and Low Pressure (LP) steams for the ST and process users. The cooled GT flue gas exits the HRSG at 270°F and is vented to the atmosphere through a stack. Two HRSG trains (2 x 50%), in tandem with the combustion turbines, are in operation.

The IGCC uses one steam turbine to generate power from the steam generated in the HRSG and process waste heat recovery (WHR) systems. HP steam at 1,800 psig and 1,000 °F and IP steam at 467 psia and 1,000 °F are used in the HP and IP stages of the ST for power generation. LP exhaust steam from the last ST stage is condensed by splitting 50/50 to a surface condenser and an air-cooled condenser to conserve cooling water. The condensers operate at 0.698 psia with a corresponding condensing temperature of 90 °F.

The condensates are collected and sent to a deaerator to remove dissolve gases and treated to provide BFW for the steam generators. Two 50% capacity BFW pumps are provided for each of the three (HP, IP and LP) steam generators.

3.2.9 Cooling Water Systems

Exhaust steam from the ST is split 50/50 to a surface condenser cooled with cooling water and to an air-cooled condenser using ambient air and forced convection. The major impact of utilizing this parallel cooling method is a significant reduction in water requirement when compared to a wet cooling system.

The circulating water system is a closed-cycle cooling water system that supplies cooling water to the surface condenser to condense one-half of the main turbine exhaust steam. The system also supplies cooling water to the AGR plant as required, and to the auxiliary cooling system. The auxiliary cooling system is a closed loop process that utilizes a higher quality water to remove heat from compressor intercoolers, oil coolers and other ancillary equipment and transfers that heat to the main circulating cooling water system in plate-and-frame heat exchangers. The heat transferred to the circulating water in the surface condenser and other applications is removed by a mechanical draft cooling tower.

The system consists of two 50% capacity vertical circulating water pumps (CWPs), a mechanical draft evaporative cooling tower, and CS cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counter-flow mechanical draft cooling tower.

3.2.10 BFW/Condensate System

The function of the BFW system is to pump the various BFW streams from the deaerator storage tank in the HRSG to the respective steam drums. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow. The BFW pumps are supplied with instrumentation to monitor and alarm on low oil pressure, or high bearing temperature. BFW pump suction pressure and temperature are also monitored. In addition, the suction of each BFW pump is equipped with a startup strainer.

3.2.11 Accessory Electric Plant

The accessory electric plant consists of switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, and wire and cable. It also includes the main transformer, all required foundations, and standby equipment.

3.2.12 Instrumentation and Control

An integrated plant-wide distributed control system (DCS) is provided, per the description given in the NETL Report 1399.

3.3 CASE 2a PROCESS DESCRIPTION

Case 2a is modeled upon the process information provided in Case S1B within the NETL Report 1399. The system description follows the BFD in Figure 3-2 and stream numbers reference the

same figure. Table 3-1 provides Nexant's model generated process data for the numbered streams in the BFD.

Coal Preparation and Drying

Coal receiving and handling is part of the IGCC common areas and covered in Section 3.2.1. Coal is crushed in the coal mill and delivered to a surge hopper, which in turn delivers the coal to the coal pre-heater. The WTA coal drying process, described in Section 3.2.2 reduces the PRB coal moisture content from 25.77 wt% to 6 wt%.

Coal Feed System

The dried coal is drawn from the surge hoppers and fed through a pressurization lock hopper system to a dense phase pneumatic conveyor, which uses nitrogen from the ASU to convey the coal to the gasifiers.

Air Separation Unit

The ASU plant is designed with two production trains. The air compressor is powered by an electric motor. Nitrogen is also recovered, compressed, and used as diluent in the GT combustor and as a coal transport fluid.

Shell Gasifier

There are two Shell dry feed, pressurized, up flow, entrained, slagging gasifiers, operating at 615 psia. Coal reacts with oxygen in a reducing environment to produce principally hydrogen and carbon monoxide with some CO₂. High-temperature heat recovery in each gasifier train is accomplished in three steps, including the gasifier membrane wall, which maintains a protective ash layer over the membrane wall. The product gas from the gasifier is cooled using a syngas recycle quench to lower the temperature below the ash melting point. Syngas then goes through a raw gas cooler, which lowers the gas temperature and contributes to the production of HP steam for use in steam cycle.

The solids are removed as both slag and ash. Liquid slag is solidified in a water bath and removed via a lock hopper system. Ash carried over with the syngas is removed in a ceramic candle filter. The collected ash is also removed from the filter via a lock hopper system. The syngas scrubber downstream of the gasifier removes any remaining Particulate Matter (PM) passing the candle filter because of leakage around the filter seals or any undetected candle failure that allows the passage of large particulates.

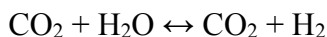
Syngas Scrubber/Sour Water Stripper

The raw syngas exiting the ceramic particulate filter enters the scrubber for removal of chlorides and any remaining particulates. The quench scrubber washes the syngas in a countercurrent flow in two packed beds, which removes essentially all traces of entrained particles. The bottoms from the scrubber are sent to the slag removal and handling system for processing.

The sour water stripper removes ammonia, sulfur, and other impurities from the waste stream of the scrubber. The sour gas stripper consists of a sour drum that accumulates sour water from the gas scrubber and condensate from low temperature syngas cooling.

Water Gas Shift and COS Hydrolysis

To enable 90% carbon capture from the syngas, the gasifier syngas product must be converted to hydrogen-rich syngas. This is achieved by converting most of the syngas CO to hydrogen and CO₂ by reacting the CO with water over a bed of catalyst, per the following reaction:



Steam for the WGS is provided partly by the vaporization of water during syngas quench cooling and partly by saturation of the overhead gas in the syngas scrubber. Additional steam is injected to the syngas stream to push the equilibrium towards a high conversion of CO. The shift catalyst is sulfur-tolerant and it also hydrolyses the COS to H₂S for removal in the AGR system.

Cooling is provided between reactors to control the exothermic temperature rise. A parallel set of reactors is required due to the high mass flow rate. The heat exchanger after the second WGS reactor is a gas-gas exchanger used to preheat the syngas prior to the first WGS reactor.

Mercury Removal and AGR

Mercury removal is achieved via an activated carbon process described in Section 3.2.4. The AGR process is a two-stage Selexol™ process where H₂S is removed in the first stage and CO₂ in the second stage of absorption. The process results in three product streams, the clean syngas, a CO₂-rich stream and an acid gas feed to the Claus plant. The acid gas contains about 17 percent H₂S and 66 percent CO₂ with a balance of primarily hydrogen (H₂).

CO₂ Compression and Dehydration

CO₂ from the AGR process is generated at 17 and 150 psia. The lower pressure CO₂ stream is compressed from 17 psia to 150 psia and then combined with the CO₂ stream at 150 psia. The combined CO₂ stream is further compressed to a supercritical condition at 2,215 psia using a multiple-stage, intercooled compressor. During compression, the CO₂ stream is dehydrated to a dew point of -40°F using a thermal swing adsorptive dryer. The raw CO₂ stream from the Selexol™ process contains over 99 percent CO₂. The dehydrated CO₂ is transported to the plant fence line for sequestration outside the battery limit (OSBL).

Claus Unit

The sulfur recovery unit (SRU) is a Claus bypass type SRU utilizing oxygen instead of air. The Claus plant produces molten sulfur by converting approximately one third of the H₂S in the feed to SO₂, then reacting the H₂S and SO₂ to sulfur and water. The combined Claus technology and tail gas recycle results in an overall sulfur recovery exceeding 99 percent. Utilizing oxygen

instead of air in the Claus plant reduces the overall cost of the sulfur recovery plant. The acid gas feed consists of acid gas from both the AGR and a vent stream from the sour water stripper in the gasifier section.

Power Block

Clean syngas from the two-stage Selexol™ unit is reheated to 385°F. It is then diluted with nitrogen to meet the heating value specifications before it enters the GT combustor. The exhaust gas exits the GT at a nominal 1,050°F and enters the HRSG where additional heat is recovered. The flue gas exits the HRSG at 270°F and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced commercially available ST using a nominal 1,800 psig/1,000°F/1,000°F steam cycle. There is no air integration between the GT and the ASU.

Balance of Plant

Balance of plant items are covered in Sections 3.2.9 through 3.2.12.

3.4 SPARING PHILOSOPHY

The sparing philosophy for Case 2a is as provided below. Single train designs are utilized throughout with exceptions where equipment has exceeded its maximum capacity limit and thus requires an additional train. The design has:

- Two ASU trains (2 x 50%)
- Two trains of coal drying (2 x 50%)
- Two trains of gasification, including dry feed system, Shell gasifier, syngas cooler, cyclone and candle filter (2 x 50%)
- Two syngas clean-up trains (2 x 50%)
- Two trains of two-stage Selexol™ (2 x 50%)
- One Claus plant for sulfur recovery (1 x 100%)
- Two gas turbines/HRSG tandems (2 x 50%)
- One steam turbine (1 x 100%)

Figure 3-2

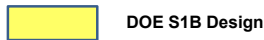


Table 3-2
Case 2a Stream Table

Description	Air to ASU	ASU Vent Gas	Sulfur Plant Oxidant	Nitrogen Diluent	Transport Nitrogen	Gasifier Oxidant	Wet Coal	Coal Moisture	Dried Coal	Slag	Hot Syngas	Syngas Quench Recycle	Quench Water	Cooled Syngas	Shift Steam	Syngas to Shift
Stream No.	1	2	3	4	5	6	8	9	10	11	12	13	14	15	16	17
V-L Mole Fraction																
Ar	0.0093	0.0165	0.0316	0.0023	0.0023	0.0318	0	0	0	0	0.0093	0.0054	0	0.0054	0	0.0046
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0	0	0.6010	0.3466	0	0.3466	0	0.2981
CO2	0.0003	0.0047	0	0	0	0	0	0	0	0	0.0327	0.0188	0	0.0188	0	0.0162
COS	0	0	0	0	0	0	0	0	0	0	0.0005	0.0003	0	0.0003	0	0.0003
H2	0	0	0	0	0	0	0	0	0	0	0.2609	0.1504	0	0.1504	0	0.1294
H2O	0.0064	0.0988	0	0	0	0	0	1.0000	0	0	0.0287	0.4391	0.9980	0.4391	1.0000	0.5198
H2S	0	0	0	0	0	0	0	0	0	0	0.0030	0.0017	0	0.0017	0	0.0015
N2	0.7759	0.7586	0.0177	0.9921	0.9921	0.0178	0	0	0	0	0.0609	0.0351	0	0.0351	0	0.0302
NH3	0	0	0	0	0	0	0	0	0	0	0.0029	0.0025	0.0020	0.0025	0	0
O2	0.2081	0.1213	0.9506	0.0054	0.0054	0.9504	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0	1.0000	0	0	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	54448	3446	202	37684	2061	11054	0	6841	0	0	38321	14737	28138	66459	9745	77266
V-L Flowrate, lb/hr	1573240	95747	6499	1057430	57841	355725	0	123241	0	0	826852	295746	506858	1333710	175559	1528540
Solids Flow Rate, lb/hr	0	0	0	0	0	0	585971	0	462730	49444	0	0	0	0	0	0
Temperature, F	42	64	90	385	387	292	42	75	158	2602	2602	474	338	450	550	450
Pressure, psia	13.0	16.4	125.0	384.0	815.0	740.0	13.0	48.0	14.8	615.0	615.0	615.0	115.0	570.0	600.0	555.0
Enthalpy, Btu/lb	-33.3	-463.2	1.8	75.1	74.8	43.8		-6823			-798	-3120	-6549	-3129	-5611	-3457
Density, lb/cuft	0.070	0.089	0.687	1.180	2.473	2.961		62.273			0.401	1.266	56.011	1.206	1.144	1.171

Case 2a Stream Table (cont'd)

Description	Syngas to Hg Removal	Tail Gas Recycle	Syngas to Selexol™	H2S to Claus	Sulfur Product	Selexol™ CO2 Out	CO2 Product	Treated Syngas	Reheat Syngas	Ambient Air to GT	GT Flue Gas	Stack Gas	HP SH Steam	Air-Cooled LP Cond	Water-Cooled LP Cond	BFW to HRSG
Stream No.	18	19	20	21	22	24	25	26	27	28	29	30	31	32	33	34
V-L Mole Fraction																
Ar	0.0060	0.0095	0.0060	0.0024	0	0.0002	0.0002	0.0095	0.0095	0.0093	0.0089	0.0089	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO	0.0099	0.0033	0.0098	0.0044	0	0.0002	0.0002	0.0156	0.0156	0	0	0	0	0	0	0
CO2	0.3974	0.6415	0.4008	0.6579	0	0.9919	0.9948	0.0491	0.0491	0.0003	0.0089	0.0089	0	0	0	0
COS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0.5437	0.1156	0.5376	0.1177	0	0.0046	0.0046	0.8618	0.8608	0	0	0	0	0	0	0
H2O	0.0018	0.1393	0.0037	0.0430	0	0.0029	0.0000	0.0001	0.0006	0.0064	0.1207	0.1207	1.0000	1.0000	1.0000	1.0000
H2S	0.0022	0.0058	0.0023	0.1650	0	0	0	0	0	0	0	0	0	0	0	0
N2	0.0391	0.0851	0.0398	0.0096	0	0.0002	0.0002	0.0638	0.0644	0.7759	0.7551	0.7551	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0.2081	0.1063	0.1063	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	1.0000	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	59670	852	60522	829	132	22046	21981	37527	37570	221242	280034	280034	48880	43893	43893	83101
V-L Flowrate, lb/hr	1211450	28987	1240440	29927	4231	964151	962980	244215	245240	6392720	7695380	7695380	880585	790735	790735	1497100
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature, F	95	102	115	119		60	162	87	385	42	1049	270	999	90	90	240
Pressure, psia	479.7	5.6	479.7	23.7		149.7	2214.7	469.6	460.0	12.9	13.5	13.0	1815.0	0.7	0.7	105.0
Enthalpy, Btu/lb	-3344	-3615	-3342	-3246		-3852		-1383	-1059	-33	-249	-461	-5387	-6808	-6808	-6658
Density, lb/cuft	1.671	0.032	1.622	0.139		1.254	28.796	0.514	0.327	0.069	0.023	0.046	2.282	62.117	62.117	59.099

3.5 PERFORMANCE RESULTS

The Nexant-modeled Case 2a IGCC plant with CO₂ capture consumes 7,032 tpd PRB coal at the Montana site and produces a net output of 461 MWe with a net plant efficiency of 31.32% on a Higher Heating Value (HHV) basis. Overall performance for the Case 2a IGCC plant is summarized in Table 3-3, which includes auxiliary power requirements.

Table 3-3
Case 2a Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case 2a
Gas Turbine Power	429,973
Steam Turbine Power	224,080
TOTAL POWER, kWe	654,053
Auxiliary Load Summary, kWe	
Coal Handling	510
Coal Milling	2,730
Slag Handling	580
WTA Coal Dryer Compressor	9,370
WTA Coal Dryer Auxiliaries	620
Air Separation Unit Auxiliaries	1,003
Air Separation Unit Main Air Compressor	63,719
Oxygen Compressor	8,830
Nitrogen Compressors	33,340
CO ₂ Compressor	31,545
Boiler Feedwater Pumps	4,331
Condensate Pump	217
Quench Water Pump	760
Syngas Recycle Compressor	820
Circulating Water Pump	3,090
Ground Water Pumps	335
Cooling Tower Fans	2,015
Air Cooled Condenser Fans	2,847
Scrubber Pumps	20
Acid Gas Removal	18,391
Gas Turbine Auxiliaries	998
Steam Turbine Auxiliaries	96
Claus Plant/TGTU Auxiliaries	249
Claus Plant TG Recycle Compressor	1,517
Miscellaneous Balance of Plant	3,000
Transformer Losses	2,514
TOTAL AUXILIARIES, kWe	193,448
NET POWER, kWe	460,605
Net Plant Efficiency, % (HHV)	31.32%
Net Plant Heat Rate, Btu/kWh	10,895
CONDENSER COOLING DUTY, MMBtu/hr	1,202
CONSUMABLES	
As-Received Coal Feed, lb/hr	585,971
Thermal Input, kWt	1,470,705
Raw Water Withdrawal, gpm	3,685
Raw Water Consumption, gpm	2,971

Table 3-4 shows the carbon balance for the Case 2a IGCC plant. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not part of the carbon capture equation, but is not neglected in the balance since the model accounts for the air components throughout. Carbon leaves the plant as unburned carbon in the slag, CO₂ in the stack gas, ASU vent gas, sour water to be treated, and the CO₂ capture product. The carbon capture efficiency is defined as the amount of carbon in the CO₂ product stream relative to the amount of carbon in the coal less carbon contained in the slag. For Case 2a, the carbon capture efficiency is 90.0%.

Table 3-4
Case 2a Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	293,384	
ASU Air	196	
Air to Gas Turbine	797	
ASU Vent		196
Carbon in Slag		1,467
Sulfur Product		0
Stack Gas		29,973
CO ₂ Product		262,741
Convergence Tolerance		0
Total	294,377	294,377

Table 3-5 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur emitted in the stack gas, and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible.

Table 3-5
Case 2a Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	4,263	
Sulfur Product		4,253
Stack Gas		2
CO ₂ Product		8
Convergence Tolerance		0
Total	4,263	4,263

Table 3-6 shows the overall water balance for the Case 2a IGCC plant. Raw water is obtained from groundwater (50%) and from municipal sources (50%). Water demand represents the total

amount of water required for a particular process. Some water is recovered within the process, primarily from the coal drying process and as syngas condensate, and that water is used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. Water is discharged from the overall process in the cooling tower and with the slag. The net water consumption is the difference between the raw water withdrawal and process water discharge.

Table 3-6
Case 2a Overall Water Balance

Water Use, gpm	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Net Water Consumption
Slag Handling	124	124	0	0	0
Quench/Wash	1,013	755	258	0	258
SWS Blowdown	0	0	0	0	0
Condenser Makeup	535	0	535	0	535
Shift Steam	351	0	351		
BFW Makeup	15	0	15		
Scrubber Makeup	169	0	169		
Claus Plant	0	0	0		0
Cooling Tower	3,151	259	2,892	714	2,178
Water from Coal Drying		246	-246		
BFW Blowdown		13	-13		
SWS Blowdown		0	0		
Total	4,823	1,138	3,685	714	2971

3.6 EQUIPMENT LIST

As the Case 2a Shell IGCC is based on Case S1B in NETL Report 1399, the reader should refer to the Case S1B equipment list in NETL Report 1399.

3.7 CAPITAL COST

Table 3-7 shows the cost breakdown of the Case 2a Shell Gasifier with Selexol™-based AGR IGCC, consistent with the code of accounts format as expressed in the NETL Report 1399. The accounts/subaccounts of interest for this study are:

- 2.3 Dry coal injection system,
- 4 Gasifier & accessories,
- 5A Gas cleanup & piping and
- 5B CO₂ removal and compression.

These are shown with more detail to include the various subaccounts and provide more clarity to the major cost differences among the cases.

Table 3-8 shows the calculation and addition of owner's costs to determine the TOC, used to calculate COE.

The estimated TOC of the Case 2a Shell Gasifier with Selexol™ AGR IGCC using PRB coal in 2011 dollars is \$5,400/kW.

Table 3-7
Case 2a Total Plant Cost Summary

Case 2a: Shell IGCC with Selexol™-Based AGR Low Rank Western Coal Baseline Study			PRB	Coal Feed, lb/hr Coal HHV, Btu/lb		585,971 8,564	Plant Size Net Efficiency			460.6 MW, net 31.3%		
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost \$	Eng'g CM H.O & Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect				Process	Project	\$	\$/kW
	1 COAL & SORBENT HANDLING	\$19,442	\$3,409	\$14,872	\$0	\$0	\$37,723	\$3,423	\$0	\$8,228	\$49,375	\$107
	2 COAL & SORBENT PREP & FEED											
	2.3 Dry Coal Injection System	\$57,741	\$0	\$39,047	\$0	\$0	\$96,788	\$8,337	\$0	\$21,025	\$126,150	\$274
	2.x Other Coal & Sorbent Prep & Feed Systems	\$62,434	\$10,880	\$17,044	\$0	\$0	\$90,358	\$7,894	\$0	\$19,649	\$117,901	\$256
	SUBTOTAL 2.	\$120,176	\$10,880	\$56,091	\$0	\$0	\$187,147	\$16,230	\$0	\$40,674	\$244,051	\$530
	3 FEEDWATER & MISC BOP SYSTEMS	\$9,280	\$7,068	\$9,151	\$0	\$0	\$25,499	\$2,408	\$0	\$6,472	\$34,379	\$75
	4 GASIFIER & ACCESSORIES											
	4.1 Gasifier, Syngas Cooler & Auxiliaries (Shell)	\$189,728	\$0	\$81,332	\$0	\$0	\$271,060	\$24,215	\$37,263	\$51,121	\$383,659	\$833
	4.3 ASU/Oxidant Compression	\$221,843	\$0	\$0	\$0	\$0	\$221,843	\$21,503	\$0	\$24,335	\$267,681	\$581
	4.4 LT Heat Recovery & FG Saturation	\$30,562	\$0	\$29,091	\$0	\$0	\$59,653	\$5,822	\$0	\$13,094	\$78,569	\$171
	4.6 Flare Stack System	\$0	\$1,721	\$696	\$0	\$0	\$2,417	\$232	\$0	\$529	\$3,179	\$7
	4.9 Gasification Foundations	\$0	\$10,698	\$6,380	\$0	\$0	\$17,078	\$1,564	\$0	\$4,660	\$23,302	\$51
	SUBTOTAL 4.	\$442,133	\$12,419	\$117,499	\$0	\$0	\$572,052	\$53,336	\$37,263	\$93,740	\$756,390	\$1,642
	5A GAS CLEANUP & PIPING											
	5A.1 Double Stage Selexol™	\$83,479	\$0	\$70,351	\$0	\$0	\$153,830	\$14,877	\$30,766	\$39,894	\$239,368	\$520
	5A.2 Elemental Sulfur Plant	\$5,124	\$998	\$6,566	\$0	\$0	\$12,688	\$1,232	\$0	\$2,784	\$16,705	\$36
	5A.3 Mercury Removal	\$3,489	\$0	\$2,637	\$0	\$0	\$6,125	\$591	\$306	\$1,404	\$8,426	\$18
	5A.4 Shift Reactors	\$8,938	\$0	\$3,574	\$0	\$0	\$12,511	\$1,199	\$0	\$2,743	\$16,453	\$36
	5A.6 Blowback Gas Systems	\$2,680	\$451	\$254	\$0	\$0	\$3,385	\$321	\$0	\$741	\$4,447	\$10
	5A.7 Fuel Gas Piping	\$0	\$1,047	\$684	\$0	\$0	\$1,731	\$161	\$0	\$379	\$2,270	\$5
	5A.9 HGCU Foundations	\$0	\$944	\$636	\$0	\$0	\$1,581	\$145	\$0	\$518	\$2,243	\$5
	SUBTOTAL 5.	\$103,709	\$3,440	\$84,701	\$0	\$0	\$191,851	\$18,528	\$31,072	\$48,462	\$289,912	\$629
	5B CO2 REMOVAL & COMPRESSION											
	5B.1 CO2 Removal System	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
	5B.2 CO2 Compression & Drying	\$37,766	\$0	\$12,616	\$0	\$0	\$50,381	\$4,850	\$0	\$11,046	\$66,278	\$144
	SUBTOTAL 5B.	\$37,766	\$0	\$12,616	\$0	\$0	\$50,381	\$4,850	\$0	\$11,046	\$66,278	\$144
	6 COMBUSTION TURBINE/ACCESSORIES	\$111,211	\$923	\$8,948	\$0	\$0	\$121,083	\$11,476	\$11,909	\$14,882	\$159,350	\$346
	7 HRSG, DUCTING & STACK	\$31,810	\$2,997	\$9,301	\$0	\$0	\$44,108	\$4,168	\$0	\$5,590	\$53,866	\$117
	8 STEAM TURBINE GENERATOR	\$75,940	\$985	\$19,237	\$0	\$0	\$96,162	\$9,214	\$0	\$17,626	\$123,002	\$267
	9 COOLING WATER SYSTEM	\$5,916	\$7,860	\$7,214	\$0	\$0	\$20,989	\$1,951	\$0	\$4,915	\$27,856	\$60
	10 ASH/SPENT SORBENT HANDLING SYS	\$23,279	\$1,783	\$11,463	\$0	\$0	\$36,525	\$3,505	\$0	\$4,372	\$44,402	\$96
	11 ACCESSORY ELECTRIC PLANT	\$36,081	\$16,053	\$29,135	\$0	\$0	\$81,269	\$6,994	\$0	\$16,927	\$105,189	\$228
	12 INSTRUMENTATION & CONTROL	\$12,908	\$2,617	\$8,487	\$0	\$0	\$24,012	\$2,175	\$1,200	\$4,595	\$31,981	\$69
	13 IMPROVEMENTS TO SITE	\$3,743	\$2,206	\$9,823	\$0	\$0	\$15,772	\$1,558	\$0	\$5,199	\$22,529	\$49
	14 BUILDINGS & STRUCTURES	\$0	\$7,728	\$8,781	\$0	\$0	\$16,510	\$1,502	\$0	\$2,959	\$20,970	\$46
CALCULATED TOTAL COST		\$1,033,395	\$80,369	\$407,318	\$0	\$0	\$1,521,081	\$141,319	\$81,444	\$285,688	\$2,029,531	\$4,406

Table 3-8
Case 2a Total Plant Cost Summary

Owner's Costs	\$ x \$1,000	\$/kW
Preproduction Costs		
6 months All Labor	\$16,801	\$36
1 Month Maintenance Materials	\$3,801	\$8
1 Month Non-Fuel Consumables	\$483	\$1
1 Month Waste Disposal	\$456	\$1
25% of 1 Months Fuel Cost at 100% CF	\$1,050	\$2
2% of TPC	\$40,591	\$88
Total	\$63,181	\$137
Inventory Capital		
60 day supply of fuel at 100% CF	\$8,282	\$18
60 day supply of non-fuel consumables at 100% CF	\$721	\$2
0.5% of TPC (spare parts)	\$10,148	\$22
Total	\$19,150	\$42
Initial Cost for Catalyst and Chemicals	\$15,343	\$33
Land	\$900	\$2
Other Owner's Cost	\$304,430	\$661
Financing Costs	\$54,797	\$119
Total Owner's Costs	\$457,802	\$994
Total Overnight Costs (TOC)	\$2,487,333	\$5,400

3.8 OPERATING COSTS

Table 3-9 shows the operating cost breakdown for the Case 2a IGCC.

Table 3-9
Case 2a Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Case 2a - Shell IGCC w/Selexol™-based AGR and CO2 Capture				
Plant Size (MWe)	461	Heat Rate (Btu/kWh):	10,895		
Primary/Secondary Fuel:	PRB	Fuel Cost (\$/MMBtu):			
Design/Construction	5 years	Book Life (yrs):	20		
TPC (Plant Cost) Year	June 2011	TPI Year:	2016		
Capacity Factor (%)	80	CO2 Captured (TPD)	11548		
OPERATING & MAINTENANCE LABOR					
Operating Labor					
Operating Labor Rate (base):	\$39.70	\$/hr			
Operating Labor Burden:	30.00	% of base			
Labor Overhead Charge	25.00	% of labor			
Operating Labor Requirements per Shift	units/mod	Total Plant			
Skilled Operator	2.0	2.0			
Operator	10.0	10.0			
Foreman	1.0	1.0			
Lab Tech's etc	3.0	3.0			
TOTAL Operating Jobs	16.0	16.0			
			Annual Cost	Annual Unit Cost	
			\$	\$/kW-net	
Annual Operating Labor Cost			\$7,233,658		
Maintenance Labor Cost			\$19,647,891		
Administration & Support Labor			\$6,720,387		
Property Taxes and Insurance			\$40,590,623		
TOTAL FIXED OPERATING COSTS			\$74,192,558		
VARIABLE OPERATING COSTS					
					\$/kWh-net
Maintenance Material Cost				\$36,488,940	
<u>Consumables</u>	<u>Consumption</u>	<u>Unit</u>	<u>Initial Fill</u>		
	<u>Initial</u>	<u>/Day</u>	<u>Cost</u>		
Water(/1000 gallons)	0	2,652	1.67	\$0	\$1,296,014
Chemicals					
MU & WT Chem (lb)	0	15797	0.27	\$0	\$1,235,509
Carbon (Hg Removal) (lb)	116627	160	1.63	\$190,102	\$76,154
COS Catalyst (m3)	0	0	3751.70	\$0	\$0
Water Gas Shift Catalyst (ft3)	6023	4.12	771.99	\$4,649,335	\$929,220
Selexol™ Solution (gal)	285508	90.96	36.79	\$10,503,835	\$977,132
SCR Catalyst (m3)	0	0	0.00	\$0	\$0
Ammonia (19% NH3) (ton)	0	0	0.00	\$0	\$0
Claus Catalyst (ft3)	w/equip	2.00	203.15	\$0	\$118,670
Subtotal Chemicals			\$15,343,273	\$3,336,684	
Other					
Supplemental Fuel (MMBtu)	0	0	0.00	\$0	\$0
Gases, N2 etc.(/100scf)	0	0	0.00	\$0	\$0
LP Steam (/1000 lbs)	0	0	0.00	\$0	\$0
Subtotal Other			\$0	\$0	
Waste Disposal:					
Spent Mercury Catalyst (lb)	0	160	0.65	\$0	\$30,368
Flyash (ton)	0	0	0.00	\$0	\$0
Slag (ton)	0	593	25.11	\$0	\$4,350,327
Subtotal Waste Disposal			\$0	\$4,380,695	
By-products & Emissions					
Sulfur (tons)	0	51	0.00	\$0	\$0
Subtotal By-Products			\$0	\$0	
TOTAL VARIABLE OPERATING COSTS			\$15,343,273	\$45,502,333	
Fuel (tons)	0	7032	19.63	\$0	\$40,305,148

3.9 COST OF ELECTRICITY

Table 3-10 shows a summary of the power output, CAPEX, operating expenditure (OPEX), COE and cost of CO₂ capture for the Case 2a Shell Gasifier with Selexol™ AGR IGCC. The Case 2a IGCC COE is estimated to be 145.3 mills/kWh

Table 3-10
Plant Performance and Economic Summary

Case	Case 2a
CAPEX, \$MM	
Total Installed Cost (TIC)	\$1,521
Total Plant Cost (TPC)	\$2,030
Total Overnight Cost (TOC)	\$2,487
OPEX, \$MM/yr (100% Capacity Factor Basis)	
Fixed Operating Cost (OC _{fix})	\$74.2
Variable Operating Cost, less Fuel (OC _{var})	\$56.9
Fuel (OC _{fuel})	\$50.4
Total OPEX	\$181.5
Power Production, MWe	
Gas Turbine	430.0
Steam Turbine	224.1
Auxiliary Power Consumption	193.4
Net Power Output	460.6
Power Generated, MWh/yr (MWh)	4,034,901
SO ₂ Emissions (lb/MWh _{gross})	0.0063
SO ₂ Emissions (lb/MMBtu _{gross})	0.0008
COE, excl CO₂ TS&M, mills/kWh	145.3
COE, incl CO₂ TS&M, mills/kWh	166.2
Cost of CO₂ Avoided excl CO₂ TS&M, \$/ton CO₂	79.8
Cost of CO₂ Avoided incl CO₂ TS&M, \$/ton CO₂	105.1

4.1 PROCESS OVERVIEW

The Case 2b IGCC power plant, like the Case 2a plant, is a Montana PRB coal-fired IGCC plant designed to generate enough hydrogen-rich fuel gas to fill two advanced GE 7F-turbines to generate a total of 430 MWe at the Montana site's elevation. To maximize power recovery, the power plant is equipped with HRSG and ST. It is designed to capture CO₂ equivalent to 90% of the raw syngas' carbon content.

The IGCC plant operates as a base-loaded unit with an annual on-stream CF of 80%.

The Case 2b AR Compact Gasifier with RTI WDP IGCC has the following characteristics that differentiate it from the Case 2a Shell Reference IGCC:

- The AR DSP feed system replaces the reference case's lockhopper system for feeding dried coal to the gasifier. Nexant had previously evaluated the AR DSP feed system in comparison to the Shell lockhopper feed system in a separate study (DE-FE0012062). The results from the earlier study were used to establish the performance and cost of the DSP feed system for this case.
- The AR compact gasifier replaces the Shell gasifier from Case 2a.
- Syngas leaving the gasifier that has been quenched enters RTI's WDP for sulfur removal at above 700°F instead of going through a low temperature scrubber per Case 2a.
- After sulfur removal in the WDP, the treated syngas in Case 2b then enters the WGS reactors. Unlike the Case 2a sour shift reactors, the Case 2b AFWGS consists of fixed-bed reactors (using commercial high-temperature sweet WGS catalyst) combined in such a manner as to significantly reduce the overall steam requirement and reactor capital cost while still meeting catalyst manufacturer's steam to CO recommendations. These reactors operate at standard temperatures for commercial high-temperature sweet WGS processes, but at a higher inlet temperature than commercial sour shift processes. Commercial sweet WGS catalysts cost about half as much as commercial sour WGS catalysts.
- After the hydrogen-rich shifted syngas is cooled, it enters an AACRP unit for CO₂ capture. Unlike Selexol™, which is a physical solvent, activated methyldiethanolamine is a chemical solvent. As ~99.9% of the sulfur compounds have been removed upstream by the WDP process, the AACRP process only has to remove CO₂ and is less complicated than the two-stage Selexol™ process. However, the AACRP unit also captures ~99% of any residual sulfur left in the syngas following WDP along with the CO₂, resulting in an overall system total sulfur reduction of >99.99% (sub-ppmv total sulfur in the final cleaned syngas).

Figure 4-1
Case 2b: AR Compact Gasifier with RTI WDP IGCC - Simplified BFD

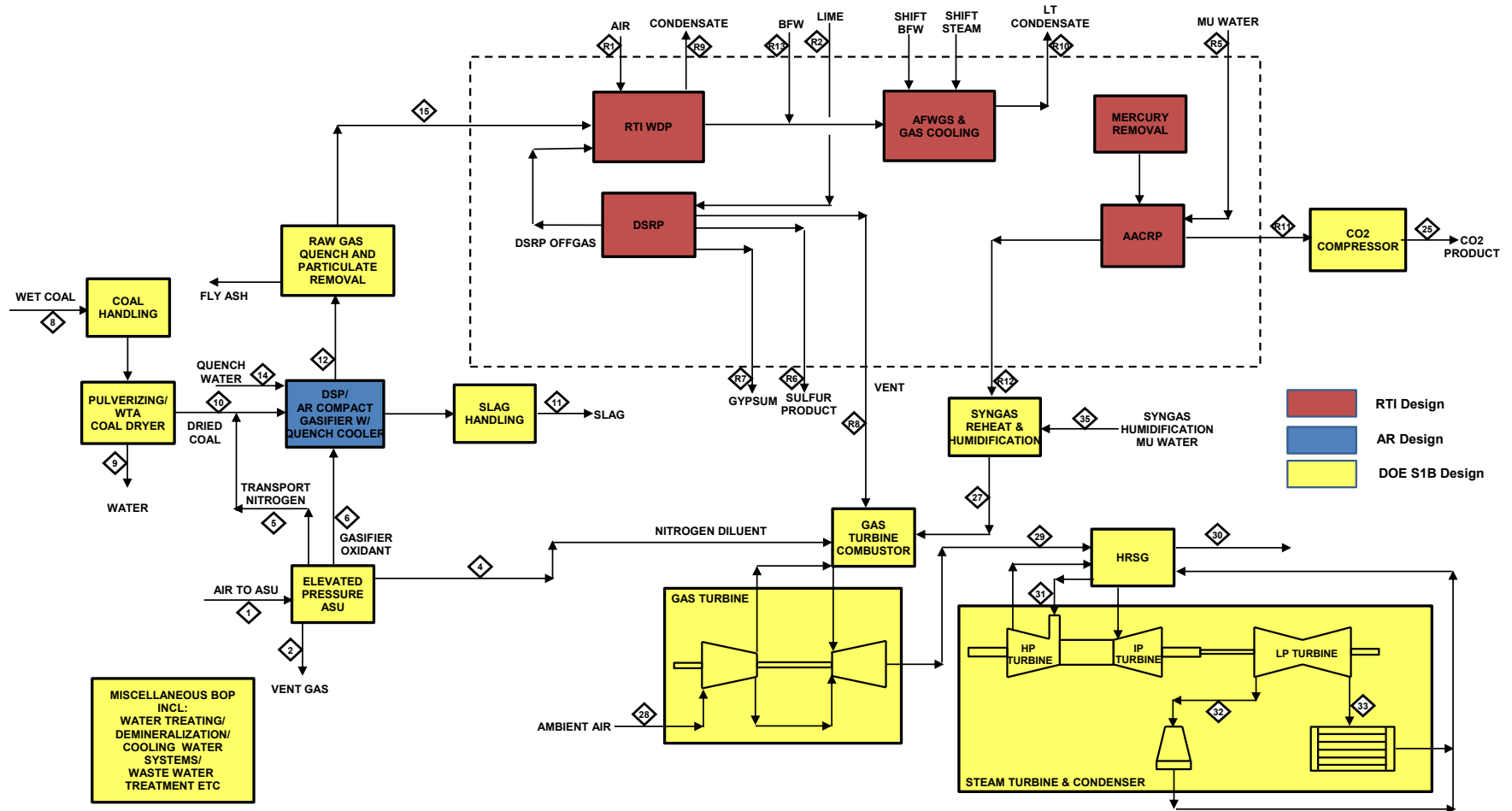


Table 4-1
Case 2b Stream Table

Description	Air to ASU	ASU Vent Gas	Nitrogen Diluent	Transport Nitrogen	Gasifier Oxidant	Wet Coal	Coal Moisture	Dried Coal	Slag	Hot Syngas	Quench Water	Cooled Syngas
Stream No.	1	2	4	5	6	8	9	10	11	12	14	15
V-L Mole Fraction												
Ar	0.0093	0.0165	0.0022	0.0023	0.0318	0	0	0	0	0.0087	0	0.0053
CH4	0	0	0	0	0	0	0	0	0	0.0087	0	0.0053
CO	0	0	0	0	0	0	0	0	0	0.6515	0	0.3954
CO2	0.0003	0.0047	0	0	0	0	0	0	0	0.0054	0	0.0033
COS	0	0	0.0	0	0	0	0	0	0	0.0004	0	0.0002
H2	0	0	0	0	0	0	0	0	0	0.2814	0	0.1708
H2O	0.0064	0.0988	0.0527	0	0	0	1.0000	0	0	0.0068	0.9981	0.3965
H2S	0	0	0	0	0	0	0	0	0	0.0033	0	0.0020
N2	0.7759	0.7586	0.9398	0.9921	0.0178	0	0	0	0	0.0323	0	0.0196
NH3	0	0	0	0	0	0	0	0	0	0.0009	0.0019	0.0013
O2	0.2081	0.1213	0.0051	0.0054	0.9504	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0	1.0000	0	0	1.0000	1.000	1.0000
V-L Flowrate, lbmol/hr	46199	2924	34692	858	9551	0	6569	0	0	35193	22797	57990
V-L Flowrate, lb/hr	1334910	81243	955052	24076	307353	0	118344	0	0	729703	410652	1140350
Solids Flow Rate, lb/hr	0	0	0	0	0	562689	0	444345	46071	0	0	0
Temperature, F	42	64	386	387	292		75			2347	290	760
Pressure, psia	13.0	16.4	384.0	815.0	740.0	13.0	48.0	14.8	615.0	615.0	115.0	605.0
Enthalpy, Btu/lb	-33.3	-463.2	-121.3	74.8	43.8		-6823			-741	-6599	-2830
Density, lb/cuft	0.070	0.089	1.159	2.473	2.961		62.273			0.420	57.620	0.910

Case 2b Stream Table (cont'd)

Description	WGD Air	Lime	MDEA MU Water	Sulfur Product	Gypsum	DSRP Vent	WDP Cond	Syngas Cool Cond	MDEA CO2 Out	Treated Syngas	Syngas Cool BFW
Stream No.	R1	R2	R5	R6	R7	R8	R9	R10	R11	R12	R13
V-L Mole Fraction											
Ar	0.0093	0	0	0	0	0.0141	0	0	0.0000	0.0088	0
CH4	0	0	0	0	0	0	0	0	0.0001	0.0087	0
CO	0	0	0	0	0	0	0	0	0.0001	0.0561	0
CO2	0.0003	0	0	0	0.0039	0.0195	0	0	0.9409	0.0010	0
COS	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0.0020	0.8853	0
H2O	0.0064	0.973695	1.0000	0	0.9759	0.0005	1.0000	0.9966	0.0569	0.0035	1.0000
H2S	0	0	0	0.0	0	0	0	0	0	0.0000	0
N2	0.7759	0	0	0	0.0003	0.9659	0	0	0.0000	0.0366	0
NH3	0	0	0	0	0	0	0	0	0	0	0
O2	0.2081	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	1.0000	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0.026305	0	0	0.0010	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0.0189	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	1138	39493	1293	105	1185	784	3	23854	22470	34541	2830
V-L Flowrate, lb/hr	32882	794001	23296	3369	24978	22322	57	429761	953647	172039	50983
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0
Temperature, F	42	75	122	600	123	100	80	100	122	122	453
Pressure, psia	13.0	13.0	511.0	447.0	30	448.0	13.0	515.0	30.7	511.0	763.0
Enthalpy, Btu/lb	-33	-7	-6817	2685		-115	-6863	-6824	-3884	-637	-6436
Density, lb/cuft	0.070	0.005	60.524	113.093		2.143	61.590	61.182	0.212	0.401	47.278

Case 2b Stream Table (cont'd)

Description	CO2 Product	Reheat Syngas	Ambient Air to GT	GT Flue Gas	Stack Gas	HP SH Steam	Air-Cooled LP Cond	Water-Cooled LP Cond	BFW to HRSG	Syngas Humid MU H2O
Stream No.	25	27	28	29	30	31	32	33	34	35
V-L Mole Fraction										
Ar	0.0000	0.0086	0.0093	0.0089	0.0089	0	0	0	0	0
CH4	0.0001	0.0085	0	0	0	0	0	0	0	0
CO	0.0001	0.0549	0	0	0	0	0	0	0	0
CO2	0.9976	0.0010	0.0003	0.0085	0.0085	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	0	0
H2	0.0021	0.8655	0	0	0	0	0	0	0	0
H2O	0	0.0258	0.0064	0.1281	0.1281	1.0000	1.0000	1.0000	1.0000	1.0000
H2S	0	0	0	0	0	0	0	0	0	0
N2	0.0000	0.0358	0.7759	0.7480	0.7480	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0
O2	0	0	0.2081	0.1065	0.1065	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	21192	35323	221242	275784	275784	56501	47147	47147	93001	2610
V-L Flowrate, lb/hr	930634	186202	6392720	7556290	7556290	1017890	849365	849365	1675430	47024
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0
Temperature, F	162	207	42	1074	270	999	90	90	220	42
Pressure, psia	2214.7	511.0	12.9	13.5	13.0	1815.0	0.7	0.7	105.0	13.0
Enthalpy, Btu/lb		-913	-33	-268	-489	-5387	-6808	-6808	-6678	-6905
Density, lb/cuft	28.796	0.371	0.069	0.022	0.046	2.282	62.117	62.117	59.628	63.210

-
- The DSRP replaces the Claus process in Case 2a. In the DSRP, sulfur leaving the WDP process in the form of SO₂ is reduced by a slipstream of shifted, hydrogen-rich syngas, forming elemental sulfur, H₂S, and COS. The elemental sulfur is condensed while the remaining H₂S and COS are re-oxidized in the presence of air to SO₂. The SO₂ is then removed in a lime scrubber, forming gypsum (CaSO₄·2H₂O).

4.2 CASE 2b PROCESS DESCRIPTION

Case 2b is modeled by adapting the Case 2a model with gasifier and syngas cleanup process information provided by AR and RTI. The system description below follows the BFD in Figure 4-1 and stream numbers referenced in the same figure. Table 4-1 provides the generated process data for the numbered streams in the BFD.

Coal Preparation and Drying

Same as Case 2a

AR DSP Coal Feed System

Dried coal from the atmospheric storage silo enters the AR DSPs via gravity flow. Three DSPs, each with a nominal capacity of 1,000 tons per day (tpd), are required to service each gasification train. The DSPs increase the pressure of the coal from atmospheric to 700 psia and subsequently discharge the coal continuously to a pressurized feed bin. The DSP also uses some nitrogen as an educting gas, which serves to evacuate and clean out the pump of solids during operation.

Coal is continuously withdrawn from the pressurized feed bin and conveyed by HP nitrogen via a single feed line to the gasifier. To maximize conversion efficiency of fuel to syngas in the gasifier, AR uses its proprietary static splitter system.

Air Separation Unit

Same as Case 2a except no oxygen is routed to the SRU.

AR Compact Gasifier

For the Case 2b IGCC, two AR compact gasifier trains operating at 615 psia are needed to generate the required amount of syngas. Fuel feeds from the pressurized feed bin via the dense phase feed line, conveyed by HP nitrogen. To maximize conversion efficiency of fuel to syngas, AR splits the feedstock from a single feed line into many small injection ports via its proprietary static splitter system. The injection ports maximize mixing of coal and oxygen to initiate the gasification reaction.

The AR compact gasifier is oriented in a vertical, down-firing position. The gasifier reaction is initiated with a torch burner, which is ignited at full gasifier pressure. The ignition torch runs on natural gas and oxygen.

The gasifier injector faceplate and the gasifier liner are water cooled to maintain the metal components at temperatures conducive to long life. The cooling water shall be clean enough (HP BFW quality) to prevent scale buildup or clogging of internal cooling passages.

The gasifier's raw syngas product is partially quenched from about 2,350°F to 760°F through the introduction of quench water spray. The quench water enters the gasifier through multiple hydraulic atomizing spray nozzles.

The solids are removed as slag and ash. Liquid slag is solidified in a water bath and removed via a lock hopper system. Ash carried over with the syngas is removed in a candle filter. The ash is also removed via a lock hopper system.

Syngas Scrubber

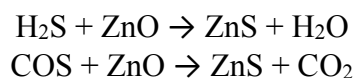
Unlike in Case 2a, the syngas scrubber is eliminated in Case 2b as the hot syngas at around 760°F enters the RTI WDP directly without requiring cooling.

Warm Syngas Cleanup Process

The Case 2b RTI advanced syngas cleanup process consists of the following system components: WDP, DSRP, AFWGS, and LTGC. AACRP is integrated with the RTI advanced syngas cleanup process for CO₂ removal. These are described in greater detail below.

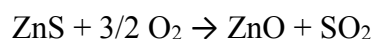
Warm Gas Desulfurization Process

The WDP process, shown in Figure 4-2, uses transport-bed reactors that are similar to commercial FCC reactor designs. It consists of a pair of these reactors: an adsorber and a regenerator. Hot syngas leaving the candle filter is routed to the WDP adsorber where it is contacted with a circulating Zn-containing attrition-resistant sorbent (developed and patented by RTI) to remove the sulfur bearing compounds, primarily H₂S and COS, from the syngas. The following reactions take place when the sorbent contacts the raw syngas:



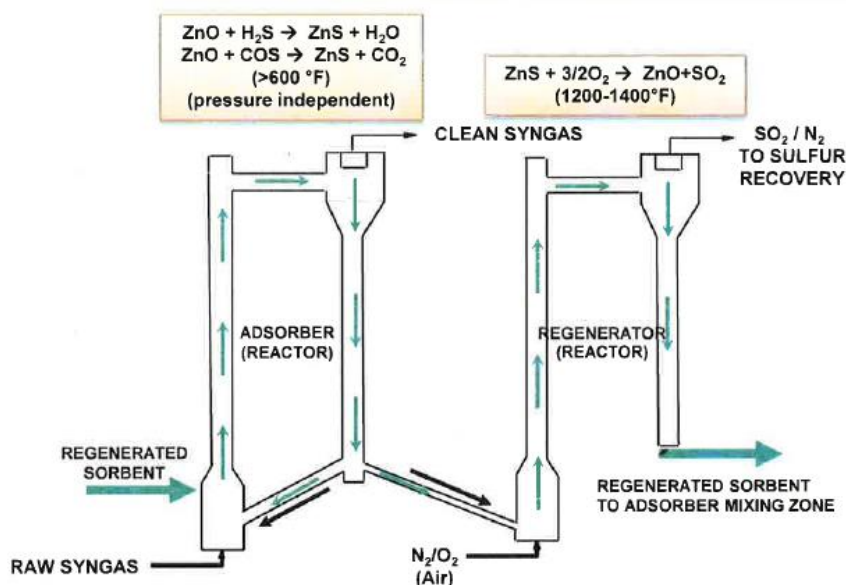
Regenerated sorbent from the regenerator, along with recycled sorbent from the adsorber standpipe, contact the raw syngas, which enters the adsorber near the bottom of the unit. The treated, essentially sulfur-free syngas is separated from sorbent via a cyclone. Any remnant attrited or fine particulate solids entrained in the essentially sulfur-free syngas are removed in a filter. A majority of the sorbent separated by the cyclone is recycled to the adsorber via a standpipe, while a portion of the sorbent is fed to the regenerator.

Within the regenerator, oxidation of the ZnS-containing sorbent takes place, producing SO₂ and regenerating ZnO, per the following reaction:



The regenerator uses air as the oxidant. Air is compressed in a multi-stage air compressor up to the regenerator operating pressure before it is fed into the regenerator. The oxidation reaction is exothermic, raising the temperature of the resulting mixture. The regenerator offgas containing SO₂ is heat exchanged with the compressed air stream before the offgas enters the DSRP. The regenerated sorbent is recycled back to the adsorber, where it adsorbs H₂S and COS again.

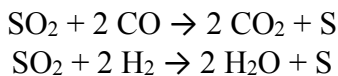
Figure 4-2
WDP Process Schematic



[Similar to Commercial FCC Reactor Designs]

Direct Sulfur Recovery Process

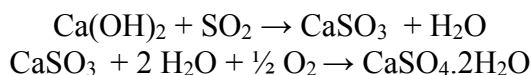
The offgas from the regenerator contains essentially SO₂ and N₂. It goes through a filter to remove any entrained solids and is cooled before it is sent into the fixed-bed DSRP reactor where SO₂ is reduced to elemental sulfur according to the following reactions:



The reducing gas is provided by a hydrogen-rich syngas slip stream from the shift reactors. A slight excess of the reducing gas is used to ensure complete reduction of the SO₂. Some H₂S and COS are formed alongside the elemental sulfur. The product stream from the DSRP reactor is sent onto a sulfur condenser unit where the elemental sulfur is condensed and separated. Heat is recovered in the condenser by making low pressure steam.

The condenser overhead gas still contains some residual H₂S and COS. These are re-oxidized to SO₂ in a fixed-bed oxidation reactor containing a redox catalyst in the presence of compressed air, which functions as the oxidant. The compressed air is a slipstream drawn from the air compressor in the WDP section.

Finally, the SO₂-containing gas leaving the oxidation reactor is cooled and sent to a lime scrubber downstream. Lime, or calcium hydroxide (Ca(OH)₂), reacts with the SO₂ in the presence of oxygen to form gypsum (CaSO₄·2H₂O) per the following reactions, which is akin to the Wet Flue Gas Desulfurization (WFGD) process used for scrubbing flue gas.



The gypsum leaves the bottom of the scrubber as a byproduct of the IGCC. The treated, sulfur-free overhead gas is split into three portions. One portion is recycled to the DSRP to help control reactor temperature rise due to the exothermic heat from the DSRP reactions. The second portion is compressed and routed to the WDP process to be utilized as the stripping or fluidizing gas in the WDP adsorber and regenerator. The remainder is routed to the GT to increase its mass flow and produce more power.

Advanced Fixed-Bed Water Gas Shift

To enable 90% carbon capture from the syngas, the gasifier syngas product must be converted to hydrogen-rich syngas. This is achieved by converting most of the syngas carbon monoxide (CO) to hydrogen and CO₂ by reacting the CO with water over a catalyst bed. As the syngas has already been treated by WDP and contains low parts-per-million sulfur, high-temperature sweet WGS catalyst, which does not need to be sulfur tolerant, is used.

RTI's AFWGS system is used to achieve an overall CO to CO₂ conversion of about 91%. Part of the steam for the WGS reaction is provided by the vaporized quench water in the AR compact gasifier. In RTI's AFWGS, a series of fixed-bed reactors using commercial high-temperature sweet WGS catalyst are combined in such a manner that enables a significant reduction in the overall steam consumption and capital cost of the WGS process, while still meeting the manufacturer's steam to CO recommendations and achieving high CO conversion required. Reducing the steam consumed in the WGS process results in higher overall thermal efficiency of the process making incorporation of carbon capture more attractive. RTI's AFWGS technology was proven as part of the 50 MWe pre-commercial demonstration of their WDP technology.

Low Temperature Gas Cooling and Mercury Removal

The shifted syngas goes through a series of heat exchangers to generate various levels of steam and preheat boiler feed water before it finally undergoes cooling by cooling water. The cooled syngas is sent to a knockout (KO) drum where the condensate is drained. Mercury is removed from the overhead syngas leaving the KO drum via the process described in Section 3.2.4.

Activated Amine CO₂ Removal Process

The AACRP is based on the activated methyldiethanolamine process, which is marketed commercially by companies such as BASF, Shell and UOP for the removal of acid gases like H₂S and CO₂. The scrubbing agent is an aqueous alkaline amine solution. For this application, the CO₂-containing syngas is passed through a reactor that contains the alkaline amine scrubbing solution, where the bulk of the CO₂ removal takes place. An acid-base reaction occurs where the CO₂ reacts with alkaline amine and is captured in solution.

To achieve the desired extent of CO₂ capture, and to ensure that the overhead gas meets the volatile organic compound (VOC) specifications, a lean amine absorber is used to treat the syngas leaving the main alkaline amine absorber. Water enters at the top of this absorber where it contacts and scrubs the CO₂-lean syngas. This serves to remove any entrained alkaline amine droplets in the syngas.

The rich, CO₂-loaded solution is sent to a regenerator to release the absorbed CO₂. The solution is first depressurized, flashing off some CO₂ in the process and helping to reduce the overall heat of CO₂ regeneration. The low-pressure solution is then sent to a thermal regenerator, where heat is applied to release the remaining CO₂. Regenerated alkaline amine solution is recycled to the absorber and used again.

It should be noted that as an alternative, hydraulic pressure recovery turbines (HPRTs) could be utilized to recover power from the depressurization of the rich amine. This reduces the total auxiliary power consumption, thus increasing net power output at the expense of greater capital cost. A trade-off study could be performed in future to determine the effects of installing the HPRTs.

CO₂ Compression and Dehydration

CO₂ from the AACRP is generated at a single pressure of 30.7 psia. The CO₂ stream is compressed to a supercritical condition at 2,215 psia using a multiple-stage, intercooled compressor. During compression, the CO₂ stream is dehydrated to a dew point of -40°F using a thermal swing adsorptive dryer. After compression and dehydration the CO₂ product meets the sequestration specifications for injection into saline aquifers, as stated in the *CO₂ Impurity Design Parameters* QGESS document. This CO₂ product is transported to the plant fence line for sequestration OSBL.

Power Block

Clean syngas from the AACRP unit, at 122°F, is contacted with hot, BFW-grade water (220°F) that was used to cool the AR gasifier. This serves to reheat the syngas and add mass to it via vaporization of the water, maximizing power generation from the GT. The fuel gas is then diluted with nitrogen and DSRP's SO₂ scrubber overhead gas to meet the heating value specifications before it enters the GT combustor. The exhaust gas exits the GT at a nominal 1,050°F and enters the HRSG where additional heat is recovered. The flue gas exits the HRSG at 270°F and is discharged through the plant stack. The steam raised in the HRSG is used to power

an advanced commercially available steam turbine using a nominal 1,800 psig/1,000°F/1,000°F steam cycle. There is no air integration between the gas turbine and the ASU.

Balance of Plant

Balance of plant items are covered in Sections 3.2.9 through 3.2.12.

4.3 SPARING PHILOSOPHY

The sparing philosophy for Case 2b is as provided below. Single train designs were utilized throughout with exceptions where equipment had exceeded its maximum capacity limit or dimensions for acceptable transportation to the site and thus required an additional train. For AACRP, multiple trains were used for specific groups of equipment to enable meeting the transportation dimension constraints. The design has:

- Two ASU trains (2 x 50%)
- Two trains of coal drying (2 x 50%)
- Two trains of gasification, including dry feed system, AR compact gasifier, syngas quench cooler, cyclone and candle filter (2 x 50%)
- One WDP train (1 x 100%)
- One DSRP train (1 x 100%)
- One AFWGS and LTGC train (1 x 100%)
- One AACRP train (3 x 33% for rich amine absorber and LP flash drum, 1 x 100% for rest of process)
- Two gas turbines/HRSG tandems (2 x 50%)
- One steam turbine (1 x 100%)

4.4 PERFORMANCE RESULTS

The Nexant-modeled Case 2b IGCC plant with CO₂ capture consumes 6,752 tpd of PRB coal at the Montana site to produce a net output of 463 MWe at a net plant efficiency of 32.75% on a HHV basis. Overall performance for the Case 2b IGCC plant is summarized in Table 4-2, which includes auxiliary power requirements. Loads that are unique to the AR and RTI processes are shown in bold and italicized.

When integrating RTI's WDP into the Case 2b IGCC, preferred process configurations avoided cooling the syngas below the WDP's typical operating temperature of >700°F, and consequently eliminated equipment to cool and subsequently reheat the syngas to this operating temperature. Compared to the reference Case S1B, this results in the elimination of the syngas cooling scrubber and syngas interchange heat exchanger between the inlet and outlet of WDP. This results in a significant reduction in equipment costs. However, the elimination of equipment in the middle of the process also changes the pressure profile across the overall process with very limited options to achieve a final outlet pressure similar to the reference case. To complete an optimized integration of the RTI WDP, the operating pressure of the gasifier system needs to be adjusted to achieve a similar final pressure of the fully treated hydrogen-rich syngas.

Table 4-2
Case 2b Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case 2b
Gas Turbine Power	429,904
Steam Turbine Power	211,301
TOTAL POWER, kWe	641,206
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	490
Coal Milling	2,622
Slag Handling	540
WTA Coal Dryer Compressor	8,998
WTA Coal Dryer Auxiliaries	595
DSP	1,803
Air Separation Unit Auxiliaries	851
Air Separation Unit Main Air Compressor	54,067
Oxygen Compressor	7,630
Nitrogen Compressors	26,624
CO ₂ Compressor	32,295
Boiler Feedwater Pumps	4,234
Condensate Pump	233
Quench Water Pump	616
Gasifier Cooling BFW Circulating Pump	3,723
Circulating Water Pump	2,752
Ground Water Pumps	322
Cooling Tower Fans	1,794
Air Cooled Condenser Fans	2,715
AACRP	16,970
Gas Turbine Auxiliaries	998
Steam Turbine Auxiliaries	91
RTI WDP	1,983
DSRP	283
Miscellaneous Balance of Plant	3,000
Transformer Losses	2,465
TOTAL AUXILIARIES, kWe	178,691
NET POWER, kWe	462,518
Net Plant Efficiency, % (HHV)	32.75%
Net Plant Heat Rate, Btu/kWh	10,419
CONDENSER COOLING DUTY, MMBtu/hr	1,147
CONSUMABLES	
As-Received Coal Feed, lb/hr	562,689
Thermal Input, kWt	1,412,271
Raw Water Withdrawal, gpm	3,532
Raw Water Consumption, gpm	2,896

In Case 2b, the final optimization to match the final pressure of the fully treated, hydrogen-rich syngas in the reference case was not completed. The consequence of this is the operating pressure for the AACRP was significantly higher than Selexol™ in the reference case. Because a significant portion of the CO₂ removal results from the pressure reduction between operating pressure and about 30 psia, the primary auxiliary load for this process is the electricity to drive the pumps that compress the liquid from about 30 psia to operating pressure. At higher operating pressures the auxiliary load was greater, resulting in a lower net efficiency for the IGCC plant. For an estimated decrease in the auxiliary load of roughly 1 MW if the pressure drop across the system had been optimized, the estimated net efficiency would be 32.82%.

Table 4-3 shows the carbon balance for the Case 2b IGCC plant. The carbon capture efficiency is defined as the amount of carbon in the CO₂ product stream relative to the amount of carbon in the coal less carbon contained in the slag. For Case 2b, the carbon capture efficiency is 90.1%.

Table 4-3
Case 2b Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	281,727	
ASU Air	166	
Air to WDP	4	
Air to Gas Turbine	797	
ASU Vent		166
LTGC Condensate to Cooling Tower		278
Carbon in Slag		0
Stack Gas		28,278
CO ₂ Product		253,972
Convergence Tolerance		
Total	282,694	282,694

Table 4-4 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the elemental sulfur condensed in the DSRP, gypsum from the lime scrubber, sulfur emitted in the stack gas, and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible.

Table 4-4
Case 2b Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	4,094	
Gypsum		713
Sulfur Product		3,369
Stack Gas		0
CO ₂ Product		12
Convergence Tolerance		0
Total	4,094	4,094

Table 4-5 shows the overall water balance for the Case 2b IGCC plant. The raw water withdrawal is the net difference between the water demand and the internal recycle. The net water consumption is the difference between the raw water withdrawal and process water discharge.

Table 4-5
Case 2b Overall Water Balance

Water Use, gpm	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Net Water Consumption
Slag Handling	119	119	0	0	0
Quench/Wash	820	782	38	0	38
KO Drum Blowdown	0	0	0	0	0
Condenser Makeup	842	0	842	0	842
Shift Steam	451	0	451		
Shift BFW	229	0	229		
WDP Makeup BFW	102	0	102		
BFW Makeup	13	0	13		
AACRP Scrubber Makeup	47	0	47		
Humidification Makeup	94	0	94		94
Cooling Tower	2,806	248	2,558	636	1,922
Water from Coal Drying		236	-236		
BFW Blowdown		11	-11		
Total	4,681	1,149	3,532	636	2896

4.5 EQUIPMENT LIST

The equipment list for the AR/RTI processes that differ from the reference Shell Gasifier with Selexol™ AGR IGCC case are shown in Table 4-6 below.

Table 4-6
Case 2b Equipment List

ACCOUNT 2 COAL PREPARATION AND FEED
Subaccount 2.3 Dry Coal Injection System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	AR DSP Feed System	AR Proprietary dry solids pump with coal feed bins	1,000 tpd	6 DSPs	0

ACCOUNT 4 GASIFIER, ASU AND ACCESSORIES
Subaccount 4.1 Gasifier & Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	Gasifier	Vertical, down-fired gasifier with multiple fuel injection ports and including water quench	3,000 tpd	2	0
2	Circulating HP BFW Pump	Centrifugal	Proprietary	2	2
3	Makeup HP BFW Pump	Centrifugal	50 gpm @ 470 ft H ₂ O	2	2
4	Ash Filters	Candle	10,500 acfm	2	0
5	Syngas Cyclone	High efficiency	600,000 lb/hr	2	0

ACCOUNT 5A SYNGAS CLEANUP
Subaccount 5A.1 RTI WDP

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	WDP Adsorption Reactor	Proprietary Transport Reactor	22,300 acfm	1	0
2	WDP Regeneration Reactor	Proprietary Transport Reactor	625 acfm	1	0
3	Adsorption Cyclone Separator	Cyclone Separator	22,300 acfm	1	0
4	Stripper Cyclone Separator	Cyclone Separator	625 acfm	1	0
5	Adsorber Filter	Candle Filter	22,300 acfm	1	0
6	Regenerator Offgas Filter	Candle Filter	625 acfm	1	0
7	Adsorber Filter Lock Hopper + Fines Bin		Proprietary	1	0

8	Regenerator Filter Lock Hopper + Fines Bin		Proprietary	1	0
9	Sorbent Feeder Package (incl Hopper)		Proprietary	1	0
10	Regenerator Air Heat Exchanger	Shell and Tube	27,000 lb/hr	1	0
11	Recycle Syngas Compressor	Centrifugal	43 acfm @ 514 psia	1	0
12	Regenerator Air Compressor	Centrifugal	7,850 acfm @ 13 psia	1	0
13	DSRP Offgas Compressor	Centrifugal	90 acfm @ 511 psia	1	0
14	Syngas Recycle Compressor	Centrifugal	5 acfm @ 591 psia	1	0

ACCOUNT 5A SYNGAS CLEANUP
Subaccount 5A.2 RTI DSRP

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	DSRP Fixed Bed Reactor	Packed Bed Reactor	119,000 lb/hr	1	0
2	Tail Gas Oxidation Reactor	Packed Bed Reactor	120,000 lb/hr	1	0
3	DSRP Feed Gas Preheater	Shell and Tube	4.8 MMBtu/hr	1	0
4	Sulfur Condenser	Shell and Tube	3,400 lb/hr sulfur	1	0
5	Oxidation Reactor Gas Preheater	Shell and Tube	2.8 MMBtu/hr	1	0
6	Liquid Sulfur Separator	Pressure Vessel	3,400 lb/hr sulfur	1	0
7	Recycle Gas Compressor	Centrifugal	850 acfm @ 448 psia	1	0
8	Pulsing Gas Compressor	Centrifugal	Average 10 acfm @ 511 psia	1	0
9	Scrubber Recycle Cooler	Shell and Tube	18.2 MMBtu/hr	1	0
10	SO ₂ Scrubber	Tray Column	1,430 lb/hr SO ₂	1	0
11	Recycle Pump	Centrifugal	1,500 gpm @ 150 ft H ₂ O	1	0
12	Lime Slurry Pump	Centrifugal	50 gpm @ 950 ft H ₂ O	1	0
13	Gypsum Filter	Bag Filter	1,200 acfm	1	0
14	Lime Makeup Drum	Pressure Vessel	17,000 lb/hr	1	0

ACCOUNT 5A SYNGAS CLEANUP
Subaccount 5A.4 Shift Reactors

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	WGS Reactor 1	Packed Bed Reactor	Proprietary	1	0
2	WGS Reactor 2	Packed Bed Reactor	Proprietary	1	0
3	WGS Reactor 3	Packed Bed Reactor	Proprietary	1	0

ACCOUNT 5B CO₂ REMOVAL & COMPRESSION
Subaccount 5B.1 AACRP

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	Rich Amine Absorber	Packed Column	217,000 cuft/hr syngas	3	0
2	Lean Amine Absorber	Packed Column w/ Wash Trays	495,000 cuft/hr syngas	1	0
3	Amine Stripper	Packed Column	1,709,000 cuft/hr gas flow	1	0
4	LP Flash	Packed Column	1,785,000 cuft/hr gas flow	3	0
5	LP Flash Reflux Drum	Reflux Drum	1,513,000 cuft/hr CO ₂	3	0
6	Lean/Rich Amine Exchanger	Shell and Tube	520,000 lb/hr	6	0
7	Amine Stripper Reboiler	Shell and Tube	560,000 lb/hr amine	6	0
8	Lean Amine Cooler	Shell and Tube	1,734,000 lb/hr amine	2	0
9	Flash Overhead Condenser	Shell and Tube	114,000 lb/hr CO ₂	3	0
10	MP Flash Condenser	Shell and Tube	5,230 lb/hr amine	1	0
11	Lean Amine Pump	Centrifugal	2,050 gpm	3	0
12	Semi-lean Amine Pump	Centrifugal	13,430 gpm	3	1
13	Rich Amine Pump	Centrifugal	13,100 gpm	1	0
14	Flash Reflux Pump	Centrifugal	51 gpm	3	0
15	Amine Stripper Bottoms Pump	Centrifugal	6,035 gpm	1	1
16	Filter Pump	Centrifugal	670 gpm	1	0
17	MP Flash Recycle KO Drum	KO Drum	15,650 cuft/hr	1	0
18	MP Flash Drum	Reflux Drum	126,019 cuft/hr	1	0
19	Recycle Gas Compressor	Centrifugal	261 acfm @ 195 psia	1	0
20	Amine Tank	Vertical Vessel	14,700 gallons	1	0
21	Filter Package	Filter Package	670 gpm	1	1
22	Filter Package Vessel	Vessel	3,350 gallons	1	1

4.6 CAPITAL COST

Table 4-7 shows the cost breakdown of the Case 2b AR Compact Gasifier with RTI WDP IGCC, expressed in a consistent format with the code of accounts in the NETL Report 1399. The accounts/subaccounts of interest for this study are:

- 2.3 Dry Coal Injection System,
- 4 Gasifier & Accessories,
- 5A Gas Cleanup & Piping and
- 5B CO₂ Removal and Compression.

These are shown with more detail to include the various subaccounts and provide more clarity to the major cost differences among the cases.

Table 4-8 shows the calculation and addition of owner's costs to determine the TOC, used to calculate COE.

The estimated TOC, in 2011 dollars, of the Case 2b AR Compact Gasifier with RTI WDP IGCC is \$4,428/kW.

Table 4-7
Case 2b Total Plant Cost Summary

Case 2b: AR Compact Gasifier IGCC with RTI WDP Low Rank Western Coal Baseline Study		PRB	Coal Feed, lb/hr Coal HHV, Btu/lb		562,689 8564	Plant Size Net Efficiency		462.5 MW, net 32.7%				
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost \$	Eng'g CM H.O & Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect				Process	Project	\$	\$/kW
	1 COAL & SORBENT HANDLING	\$18,960	\$3,324	\$14,503	\$0	\$0	\$36,787	\$3,338	\$0	\$8,024	\$48,149	\$104
	2 COAL & SORBENT PREP & FEED											
	2.3 Dry Coal Injection System	\$53,924	\$0	\$35,407	\$0	\$0	\$89,331	\$7,694	\$17,866	\$22,978	\$137,870	\$298
	2.x Other Coal & Sorbent Prep & Feed Systems	\$60,786	\$10,592	\$16,594	\$0	\$0	\$87,972	\$7,685	\$0	\$19,131	\$114,788	\$248
	SUBTOTAL 2.	\$114,710	\$10,592	\$52,001	\$0	\$0	\$177,304	\$15,380	\$17,866	\$42,109	\$252,659	\$546
	3 FEEDWATER & MISC BOP SYSTEMS	\$9,563	\$7,520	\$9,267	\$0	\$0	\$26,350	\$2,485	\$0	\$6,636	\$35,472	\$77
	4 GASIFIER & ACCESSORIES											
	4.1 Gasifier, Syngas Cooler & Auxiliaries (AR)	\$37,796	\$18,898	\$57,067	\$0	\$0	\$113,761	\$10,163	\$28,440	\$22,855	\$175,218	\$379
	4.3 ASU/Oxidant Compression	\$197,746	\$0	\$0	\$0	\$0	\$197,746	\$19,167	\$0	\$21,691	\$238,605	\$516
	4.4 LT Heat Recovery & FG Saturation	\$23,086	\$0	\$22,098	\$0	\$0	\$45,184	\$4,410	\$0	\$9,918	\$59,512	\$129
	4.6 Flare Stack System	\$0	\$1,620	\$655	\$0	\$0	\$2,274	\$218	\$0	\$498	\$2,991	\$6
	4.9 Gasification Foundations	\$0	\$10,066	\$6,003	\$0	\$0	\$16,069	\$1,471	\$0	\$4,385	\$21,925	\$47
	SUBTOTAL 4.	\$258,628	\$30,583	\$85,822	\$0	\$0	\$375,034	\$35,430	\$28,440	\$59,347	\$498,251	\$1,077
	5A GAS CLEANUP & PIPING											
	5A.1 RTI WDP	\$23,171	\$0	\$27,805	\$0	\$0	\$50,977	\$4,930	\$10,195	\$13,220	\$79,322	\$172
	5A.2 RTI DSRP	\$6,324	\$0	\$7,000	\$0	\$0	\$13,324	\$1,294	\$2,665	\$3,457	\$20,740	\$45
	5A.3 Mercury Removal	\$3,273	\$0	\$2,474	\$0	\$0	\$5,747	\$555	\$287	\$1,317	\$7,905	\$17
	5A.4 RTI Shift Reactors	\$3,645	\$0	\$1,458	\$0	\$0	\$5,103	\$489	\$0	\$1,119	\$6,711	\$15
	5A.6 Blowback Gas Systems	\$2,620	\$441	\$248	\$0	\$0	\$3,309	\$314	\$0	\$724	\$4,347	\$9
	5A.7 Fuel Gas Piping	\$0	\$858	\$561	\$0	\$0	\$1,419	\$132	\$0	\$311	\$1,862	\$4
	5A.9 Gas Clean Up Foundations	incl w/ WDP	\$0	incl w/ WDP	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
	SUBTOTAL 5.	\$39,033	\$1,299	\$39,547	\$0	\$0	\$79,879	\$7,714	\$13,147	\$20,148	\$120,888	\$261
	5B CO2 REMOVAL & COMPRESSION											
	5B.1 RTI AACRP	\$24,855	\$0	\$20,878	\$0	\$0	\$45,733	\$4,423	\$9,147	\$11,860	\$71,163	\$154
	5B.2 CO2 Compression & Drying	\$38,555	\$0	\$12,879	\$0	\$0	\$51,434	\$4,952	\$0	\$11,277	\$67,663	\$146
	SUBTOTAL 5B.	\$63,410	\$0	\$33,757	\$0	\$0	\$97,167	\$9,375	\$9,147	\$23,137	\$138,826	\$300
	6 COMBUSTION TURBINE/ACCESSORIES	\$111,211	\$923	\$8,948	\$0	\$0	\$121,083	\$11,476	\$11,909	\$14,882	\$159,350	\$345
	7 HRSG, DUCTING & STACK	\$32,165	\$2,964	\$9,336	\$0	\$0	\$44,465	\$4,202	\$0	\$5,621	\$54,288	\$117
	8 STEAM TURBINE GENERATOR	\$74,757	\$944	\$19,221	\$0	\$0	\$94,923	\$9,073	\$0	\$17,613	\$121,609	\$263
	9 COOLING WATER SYSTEM	\$5,467	\$7,313	\$6,726	\$0	\$0	\$19,506	\$1,813	\$0	\$4,573	\$25,892	\$56
	10 ASH/SPENT SORBENT HANDLING SYS	\$22,268	\$1,715	\$10,966	\$0	\$0	\$34,949	\$3,354	\$0	\$4,185	\$42,488	\$92
	11 ACCESSORY ELECTRIC PLANT	\$35,230	\$15,501	\$28,227	\$0	\$0	\$78,959	\$6,793	\$0	\$16,415	\$102,168	\$221
	12 INSTRUMENTATION & CONTROL	\$12,771	\$2,589	\$8,397	\$0	\$0	\$23,757	\$2,152	\$1,187	\$4,546	\$31,641	\$68
	13 IMPROVEMENTS TO SITE	\$3,682	\$2,170	\$9,662	\$0	\$0	\$15,514	\$1,532	\$0	\$5,114	\$22,160	\$48
	14 BUILDINGS & STRUCTURES	\$0	\$7,560	\$8,588	\$0	\$0	\$16,148	\$1,469	\$0	\$2,897	\$20,515	\$44
	CALCULATED TOTAL COST	\$801,857	\$94,999	\$344,968	\$0	\$0	\$1,241,824	\$115,587	\$81,696	\$235,248	\$1,674,355	\$3,620

Table 4-8
Case 2b Total Plant Cost Summary

Owner's Costs	\$ x \$1,000	\$/kW
Preproduction Costs		
6 months All Labor	\$14,652	\$32
1 Month Maintenance Materials	\$3,136	\$7
1 Month Non-Fuel Consumables	\$528	\$1
1 Month Waste Disposal	\$425	\$1
25% of 1 Months Fuel Cost at 100% CF	\$1,008	\$2
2% of TPC	\$33,487	\$72
Total	\$53,236	\$115
Inventory Capital		
60 day supply of fuel at 100% CF	\$7,953	\$17
60 day supply of non-fuel consumables at 100% CF	\$530	\$1
0.5% of TPC (spare parts)	\$8,372	\$18
Total	\$16,855	\$36
Initial Cost for Catalyst and Chemicals	\$6,148	\$13
Land	\$900	\$2
Other Owner's Cost	\$251,153	\$543
Financing Costs	\$45,208	\$98
Total Owner's Costs	\$373,500	\$808
Total Overnight Costs (TOC)	\$2,047,855	\$4,428

4.7 OPERATING COSTS

Table 4-9 shows the operating cost breakdown for the Case 2b IGCC.

Table 4-9
Case 2b Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Case 2b - AR Compact Gasifier IGCC w/ RTI WDP				
Plant Size (MWe)	463	Heat Rate (Btu/kWh):		10,419	
Primary/Secondary Fuel:	PRB	Fuel Cost (\$/MMBtu):			
Design/Construction	5 years	Book Life (yrs):		20	
TPC (Plant Cost) Year	June 2011	TPI Year:		2016	
Capacity Factor (%)	80	CO2 Captured (TPD)		11165	
OPERATING & MAINTENANCE LABOR					
Operating Labor					
Operating Labor Rate (base):	\$39.70	\$/hr			
Operating Labor Burden:	30.00	% of base			
Labor Overhead Charge	25.00	% of labor			
Operating Labor Requirements per Shift		units/mod	Total Plant		
Skilled Operator		2.0	2.0		
Operator		10.0	10.0		
Foreman		1.0	1.0		
Lab Tech's etc		3.0	3.0		
TOTAL Operating Jobs		16.0	16.0		
			Annual Cost	Annual Unit Cost	
			\$	\$/kW-net	
Annual Operating Labor Cost			\$7,233,658		
Maintenance Labor Cost			\$16,209,429		
Administration & Support Labor			\$5,860,771.69		
Property Taxes and Insurance			\$33,487,097		
TOTAL FIXED OPERATING COSTS			\$62,790,955		
VARIABLE OPERATING COSTS					
Maintenance Material Cost			\$30,103,226		
			\$/kW-h-net		
Consumables	Consumption	Unit	Initial Fill		
	Initial /Day	Cost	Cost		
Water(/1000 gallons)	0 2,543	1.67	\$0	\$1,243,123	
Chemicals					
MU & WT Chem (lb)	0 15152	0.27	\$0	\$1,185,086	
Carbon (Hg Removal) (lb)	111993 154	1.63	\$182,549	\$73,128	
RTI WDP Sorbent (lb)	Proprietary Proprietary	Proprietary	\$2,459,501	\$1,541,514	
RTI DSRP Catalyst (lb)	22000 12.1	11.15	\$245,300	\$39,248	
DSRP Oxidation/Reduction Catalyst (lb)	22000 12.1	1.56	\$34,320	\$5,491	
Lime	0 41904	0.0382	\$0	\$467,550	
Water Gas Shift Catalyst (ft3)	5793 3.2	394.38	\$2,284,823	\$365,572	
Activated Amine (lb)	336647 184	2.80	\$941,894	\$150,703	
Subtotal Chemicals			\$6,148,386	\$3,828,293	
Other					
Supplemental Fuel (MMBtu)	0 0	0.00	\$0	\$0	
Gases, N2 etc.(/100scf)	0 0	0.00	\$0	\$0	
LP Steam (/1000 lbs)	0 0	0.00	\$0	\$0	
Subtotal Other			\$0	\$0	
Waste Disposal:					
Spent Mercury Catalyst (lb)	0 154	0.65	\$0	\$29,161	
Flyash (ton)	0 0	0.00	\$0	\$0	
Slag (ton)	0 553	25.11	\$0	\$4,053,550	
Subtotal Waste Disposal			\$0	\$4,082,711	
By-products & Emissions					
Sulfur (tons)	0 40	0	\$0	\$0	
Gypsum (tons)	0 300	0	\$0	\$0	
Subtotal By-Products			\$0	\$0	
TOTAL VARIABLE OPERATING COSTS			\$6,148,386	\$39,257,352	
Fuel (tons)	0 6752	19.63	\$0	\$38,703,730	

4.8 COST OF ELECTRICITY

Table 4-10 shows a summary of the power output, CAPEX, OPEX, COE and cost of CO₂ capture for the Case 2b AR Compact Gasification with RTI WDP IGCC. The COE for the Case 2b IGCC is 122.0 mills/kWh

Table 4-10
Plant Performance and Economic Summary

Case	Case 2b
CAPEX, \$MM	
Total Installed Cost (TIC)	\$1,242
Total Plant Cost (TPC)	\$1,674
Total Overnight Cost (TOC)	\$2,048
OPEX, \$MM/yr (100% Capacity Factor Basis)	
Fixed Operating Cost (OC _{fix})	\$62.8
Variable Operating Cost, less Fuel (OC _{var})	\$49.1
Fuel (OC _{fuel})	\$48.4
Total OPEX	\$160.2
Power Production, MWe	
Gas Turbine	429.9
Steam Turbine	211.3
Auxiliary Power Consumption	178.7
Net Power Output	462.5
Power Generated, MWh/yr (MWh)	4,051,626
SO ₂ Emissions (lb/MWh _{gross})	0.0013
SO ₂ Emissions (lb/MMBtu)	0.0002
COE, excl CO₂ TS&M, mills/kWh	122.0
COE, incl CO₂ TS&M, mills/kWh	142.0
Cost of CO₂ Avoided excl CO₂ TS&M, \$/ton CO₂	51.1
Cost of CO₂ Avoided incl CO₂ TS&M, \$/ton CO₂	75.2

5.1 PROCESS OVERVIEW

The Case 2c IGCC power plant, like the previous two cases, is a Montana PRB coal-fired IGCC plant designed to generate hydrogen-rich fuel gas to fill two advanced GE 7F-turbines for a total gas turbine power generation of 430 MW at the Montana site's elevation. The IGCC power plant is equipped with HRSG and ST to maximize power recovery. It is designed to capture CO₂ equivalent to 90% of the raw syngas' carbon content. The IGCC plant operates as a base-loaded unit with an annual on-stream CF of 80%.

The purpose of establishing Case 2c and similarly, Case 2d, is to evaluate respectively on a standalone basis, the performance and cost of the IGCC plant utilizing the AR advanced gasification and RTI advanced syngas cleanup and AACRP systems. These cases can then be compared with Case 2b to evaluate if there are any synergistic advantages in integrating both systems in an IGCC plant.

The Case 2c IGCC has the following characteristics that differentiate it from the Case 2a Shell Reference IGCC:

- The AR DSP feed system replaces the reference case's lockhopper system used to feed dried coal to the gasifier. Nexant had previously evaluated the AR DSP feed system in comparison to the Shell lockhopper feed system in a separate study (DE-FE0012062). The results from the earlier study are used to establish the performance and cost of the DSP feed system for this case.
- The AR compact gasifier replaces the Shell gasifier from Case 2a.

The coal feed rate is the same as Case 2b since both utilize the AR gasification process operating under the same conditions, hence same cold gas efficiency, to generate enough syngas for firing in the two GE 7F-turbines.

5.2 CASE 2c PROCESS DESCRIPTION

The system description below for Case 2c follows the BFD in Figure 5-1 and stream numbers reference the same figure. Table 5-1 provides the Nexant generated process data for the numbered streams in the BFD.

Coal Preparation and Drying

Same as Case 2a

AR DSP Coal Feed System

Same as Case 2b

Air Separation Unit

Same as Case 2a

AR Compact Gasifier

Same as Case 2b

Syngas Scrubber/Sour Water Stripper

Same as Case 2a

Water Gas Shift and COS Hydrolysis

Same as Case 2a

Mercury Removal and AGR

Same as Case 2a

CO₂ Compression and Dehydration

Same as Case 2a

Claus Unit

Same as Case 2a

Power Block

Same as Case 2b

Balance of Plant

Balance of plant items are covered in Sections 3.2.9 through 3.2.12.

Figure 5-1
Case 2c: AR Compact Gasifier with Selexol™ AGR IGCC - Simplified BFD

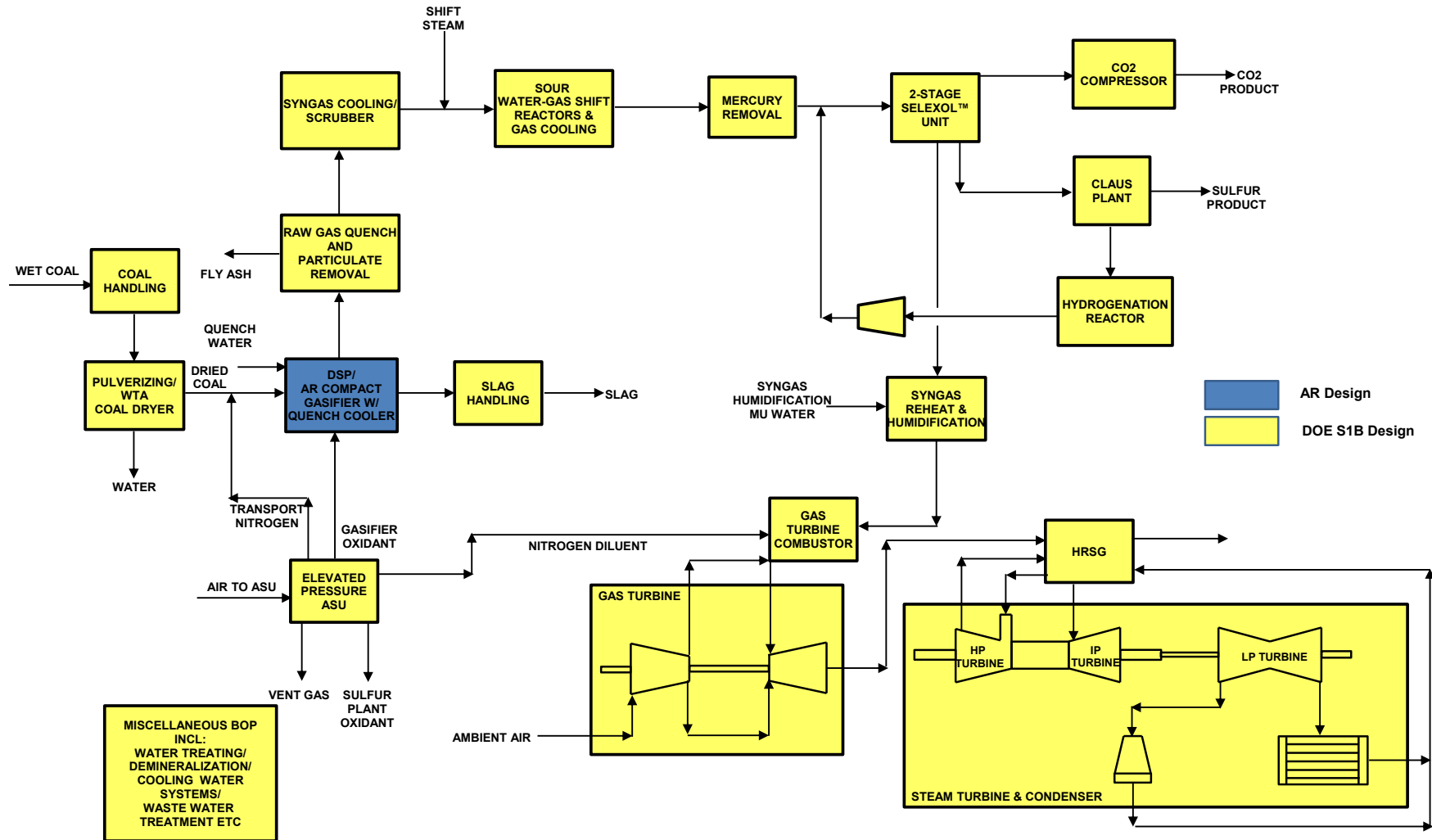


Table 5-1
Case 2c Stream Table

Description	Air to ASU	ASU Vent Gas	Sulfur Plant Oxidant	Nitrogen Diluent	Transport Nitrogen	Gasifier Oxidant	Wet Coal	Coal Moisture	Dried Coal	Slag	Hot Syngas	Recycle Syngas	Quench Water	Cooled Syngas	Shift Steam	Syngas to Shift
Stream No.	1	2	3	4	5	6	8	9	10	11	12	13	14	15	16	17
V-L Mole Fraction																
Ar	0.0093	0.0165	0.0316	0.0022	0.0023	0.0318	0	0	0	0	0.0087	0.0053	0	0.0053	0	0.0042
CH4	0	0	0	0	0	0	0	0	0	0	0.0085	0.0052	0	0.0052	0	0.0041
CO	0	0	0	0	0	0	0	0	0	0	0.6514	0.3954	0	0.3954	0	0.3118
CO2	0.0003	0.0047	0	0	0	0	0	0	0	0	0.0055	0.0033	0	0.0033	0	0.0026
COS	0	0	0	0	0	0	0	0	0	0	0.0004	0.0002	0	0.0002	0	0.0002
H2	0	0	0	0	0	0	0	0	0	0	0.2816	0.1709	0	0.1709	0	0.1348
H2O	0.0064	0.0988	0	0.049	0	0	0	1.0000	0	0	0.0069	0.3965	0.9981	0.3965	1.0000	0.5254
H2S	0	0	0	0	0	0	0	0	0	0	0.0033	0.0020	0	0.0020	0	0.0016
N2	0.7759	0.7586	0.0177	0.9434	0.9921	0.0178	0	0	0	0	0.0324	0.0196	0	0.0196	0	0.0155
NH3	0	0	0	0	0	0	0	0	0	0	0.0009	0.0013	0.0019	0.0013	0	0
O2	0.2081	0.1213	0.9506	0.0051	0.0054	0.9504	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0	1.0000	0	0	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	46853	2966	129	35063	858	9557	0	6569	0	0	35206	0	22797	58003	9966	73561
V-L Flowrate, lb/hr	1353800	82392	4155	966647	24076	307545	0	118344	0	0	729895	0	410652	1140550	179540	1420720
Solids Flow Rate, lb/hr	0	0	0	0	0	0	562689	0	444345	46071	0	0	0	0	0	0
Temperature, F	42	64	90	383	387	292	42	75	158	2349	2349	676	250	645	550	450
Pressure, psia	13.0	16.4	125.0	384.0	815.0	740.0	13.0	48.0	14.8	615.0	615.0	615.0	115.0	570.0	600.0	555.0
Enthalpy, Btu/lb	-33.3	-463.2	1.8	-108.8	74.8	43.8		-6823			-741	-2866	-6669	-2878	-5596	-3493
Density, lb/cuft	0.070	0.089	0.687	1.165	2.473	2.961		62.273			0.420	0.997	55.848	0.951	1.135	1.144

Case 2c Stream Table (cont'd)

Description	Tail Gas Recycle	Syngas to Selexol™	H2S to Claus	Sulfur Product	Selexol™ CO2 Out	CO2 Product	Treated Syngas	Reheat Syngas	Ambient Air to GT	GT Flue Gas	Stack Gas	HP SH Steam	Air-Cooled LP Cond	Water-Cooled LP Cond	BFW to HRSG	Syngas Humid MU H2O
Stream No.	19	20	21	22	24	25	26	27	28	29	30	31	32	33	34	35
V-L Mole Fraction																
Ar	0.0072	0.0054	0.0021	0	0.0002	0.0002	0.0085	0.0083	0.0093	0.0088	0.0088	0	0	0	0	0
CH4	0	0.0051	0	0	0	0	0.0083	0.0081	0	0	0	0	0	0	0	0
CO	0.0033	0.0112	0.0050	0	0.0002	0.0002	0.0178	0.0174	0	0	0	0	0	0	0	0
CO2	0.6508	0.3960	0.6542	0	0.9917	0.9948	0.0390	0.0379	0.0003	0.0087	0.0087	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0.1232	0.5559	0.1224	0	0.0047	0.0047	0.8934	0.8708	0	0	0	0	0	0	0	0
H2O	0.1537	0.0039	0.0452	0	0.0031	0.0000	0.0001	0.0255	0.0064	0.1329	0.1329	1.0000	1.0000	1.0000	1.0000	1.0000
H2S	0.0059	0.0023	0.1661	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0.0558	0.0204	0.0049	0	0.0001	0.0001	0.0328	0.0319	0.7759	0.7439	0.7439	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0.2081	0.1057	0.1057	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	1.0000	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	797	58106	791	127	21265	21200	35930	36858	221242	276792	276792	54019	44179	44179	79222	2649
V-L Flowrate, lb/hr	26959	1154810	28391	4064	929834	928661	194430	211083	6392720	7570450	7570450	973170	795895	795895	1427210	47714
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature, F	102	115	119		60	162	87	203	42	1070	270	999	90	90	242	42
Pressure, psia	5.6	473.7	23.7		149.7	2214.7	469.6	469.6	12.9	13.5	13.0	1815.0	0.7	0.7	105.0	13.0
Enthalpy, Btu/lb	-3730	-3418	-3254		-3853		-1415	-1618	-33	-289	-509	-5387	-6808	-6808	-6656	-6905
Density, lb/cuft	0.031	1.552	0.138		1.254	28.796	0.427	0.374	0.069	0.022	0.045	2.282	62.117	62.117	59.048	63.210

5.3 SPARING PHILOSOPHY

The sparing philosophy for Case 2c is as provided below. Single train designs were utilized throughout with exceptions where equipment had exceeded its maximum capacity limit or dimensions for acceptable transportation to the site and thus required an additional train. The design has:

- Two ASU trains (2 x 50%)
- Two trains of coal drying (2 x 50%)
- Two trains of gasification, including dry feed system, AR compact gasifier, syngas quench cooler, cyclone and candle filter (2 x 50%)
- Two syngas clean-up trains (2 x 50%)
- Two trains of two-stage Selexol™ (2 x 50%)
- One Claus plant for sulfur recovery (1 x 100%)
- Two gas turbines/HRSG tandems (2 x 50%)
- One steam turbine (1 x 100%)

5.4 PERFORMANCE RESULTS

The Nexant-modeled Case 2c IGCC plant with CO₂ capture consumes 6,752 tpd of PRB coal at the Montana site to produce a net output of 462 MWe at a net plant efficiency of 32.70% on a HHV basis. Overall performance for the Case 2c IGCC plant is summarized in Table 5-2, which includes auxiliary power requirements. Loads that are unique to the AR gasification system are shown in bold and italicized.

Table 5-2
Case 2c Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case 2c
Gas Turbine Power	430,245
Steam Turbine Power	209,296
TOTAL POWER, kWe	639,542
Auxiliary Load Summary, kWe	
Coal Handling	490
Coal Milling	2,622
Slag Handling	540
WTA Coal Dryer Compressor	8,998
WTA Coal Dryer Auxiliaries	595
DSP	1,803
Air Separation Unit Auxiliaries	863
Air Separation Unit Main Air Compressor	54,832
Oxygen Compressor	7,634
Nitrogen Compressors	26,951
CO ₂ Compressor	30,421
Boiler Feedwater Pumps	3,933
Condensate Pump	219
Quench Water Pump	616
Gasifier Cooling BFW Circulating Pump	3,715
Circulating Water Pump	2,801
Ground Water Pumps	320
Cooling Tower Fans	1,826
Air Cooled Condenser Fans	2,646
Scrubber Pumps	17
Acid Gas Removal	17,733
Gas Turbine Auxiliaries	998
Steam Turbine Auxiliaries	90
Claus Plant/TGTU Auxiliaries	239
Claus Plant TG Recycle Compressor	1,420
Miscellaneous Balance of Plant	3,000
Transformer Losses	2,458
TOTAL AUXILIARIES, kWe	177,778
NET POWER, kWe	461,763
Net Plant Efficiency, % (HHV)	32.70%
Net Plant Heat Rate, Btu/kWh	10,436
CONDENSER COOLING DUTY, MMBtu/hr	1,117
CONSUMABLES	
As-Received Coal Feed, lb/hr	562,689
Thermal Input, kWt	1,412,271
Raw Water Withdrawal, gpm	3517
Raw Water Consumption, gpm	2,870

Table 5-3 shows the carbon balance for the Case 2c IGCC plant. The carbon capture efficiency is defined as the amount of carbon in the CO₂ product stream relative to the amount of carbon in the coal less carbon contained in the slag. For Case 2c, the carbon capture efficiency is 90.0%.

Table 5-3
Case 2c Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	281,727	
ASU Air	169	
Air to Gas Turbine	797	
ASU Vent		169
Carbon in Slag		0
Stack Gas		28,916
CO ₂ Product		253,607
Convergence Tolerance		1
Total	282,693	282,693

Table 5-4 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur emitted in the stack gas, and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible.

Table 5-4
Case 2c Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	4,094	
Sulfur Product		4,085
Stack Gas		2
CO ₂ Product		8
Convergence Tolerance		0
Total	4,094	4,095

Table 5-5 shows the overall water balance for the Case 2c IGCC plant. The raw water withdrawal is the net difference between the water demand and the internal recycle. The net water consumption is the difference between the raw water withdrawal and process water discharge.

Table 5-5
Case 2c Overall Water Balance

Water Use, gpm	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Net Water Consumption
Slag Handling	124	124	0	0	0
Quench/Wash	820	707	113	0	113
SWS Blowdown	0	0	0	0	0
Condenser Makeup	701	0	701	0	701
Shift Steam	359	0	359		
BFW Makeup	13	0	13		
Scrubber Makeup	329	0	329		
Humidification Makeup	95	0	95		95
Claus Plant Condensate	0	0	0		0
Cooling Tower	2,856	248	2,608	647	1,961
Water from Coal Drying		236	-236		
BFW Blowdown		12	-12		
SWS Blowdown		0	0		
Total	4,596	1,079	3,517	647	2870

5.5 EQUIPMENT LIST

The list in Table 5-6 below shows the equipment that differs from the reference Shell Gasifier with Selexol™ AGR IGCC case.

Table 5-6
Case 2c Equipment List

ACCOUNT 2 COAL PREPARATION AND FEED
Subaccount 2.3 Dry Coal Injection System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	AR DSP Feed System	AR Proprietary dry solids pump with coal feed bins	1,000 tpd	6	0

ACCOUNT 4 GASIFIER, ASU AND ACCESSORIES
Subaccount 4.1 Gasifier & Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	Gasifier	Vertical, down-fired gasifier with multiple fuel injection ports and including water quench	3,000 tpd	2	0
2	Circulating HP BFW Pump	Centrifugal	Proprietary	2	2
3	Makeup HP BFW Pump	Centrifugal	50 gpm @ 470 ft H ₂ O	2	2
4	Ash Filters	Candle	10,000 acfm	2	0
5	Syngas Cyclone	High efficiency	600,000 lb/hr	2	0

5.6 CAPITAL COST

Table 5-7 shows the cost breakdown of the Case 2c AR Compact Gasifier with Selexol™ AGR IGCC, expressed in a format that is consistent with the code of accounts in the NETL Report 1399. The accounts/subaccounts of interest for this study are:

- 2.3 Dry Coal Injection System,
- 4 Gasifier & Accessories,
- 5A Gas Cleanup & Piping and
- 5B CO₂ Removal and Compression

These are shown with more detail to include the various subaccounts to provide more clarity as to where the major cost differences are when compared to the other cases.

Table 5-8 shows the calculation and addition of owner's costs to determine the TOC, used to calculate COE.

The estimated TOC of the Case 2c AR Compact Gasifier with Selexol™ AGR IGCC using PRB coal in 2011 dollars is \$4,709/kW.

Table 5-7
Case 2c Total Plant Cost Summary

Case 2c: AR Compact Gasifier IGCC with Selexol™-Based AGR				Coal Feed, lb/hr		562,689		Plant Size		461.8 MW, net			
Low Rank Western Coal Baseline Study				PRB		Coal HHV, Btu/lb		8,564		Net Efficiency		32.7%	
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost \$	Eng'g CM H.O & Fee	Contingencies		TOTAL PLANT COST		
				Direct	Indirect				Process	Project	\$	\$/kW	
1	COAL & SORBENT HANDLING	\$18,960	\$3,324	\$14,503	\$0	\$0	\$36,787	\$3,338	\$0	\$8,024	\$48,149	\$104	
2	COAL & SORBENT PREP & FEED												
2.3	Dry Coal Injection System	\$53,924	\$0	\$35,407	\$0	\$0	\$89,331	\$7,694	\$17,866	\$22,978	\$137,870	\$299	
2.x	Other Coal & Sorbent Prep & Feed Systems	\$60,786	\$10,592	\$16,594	\$0	\$0	\$87,972	\$7,685	\$0	\$19,131	\$114,788	\$249	
	SUBTOTAL 2.	\$114,710	\$10,592	\$52,001	\$0	\$0	\$177,304	\$15,380	\$17,866	\$42,109	\$252,659	\$547	
3	FEEDWATER & MISC BOP SYSTEMS	\$9,419	\$7,353	\$9,160	\$0	\$0	\$25,932	\$2,447	\$0	\$6,543	\$34,922	\$76	
4	GASIFIER & ACCESSORIES												
4.1	Gasifier, Syngas Cooler & Auxiliaries (AR)	\$37,721	\$18,860	\$56,953	\$0	\$0	\$113,534	\$10,143	\$28,384	\$22,809	\$174,869	\$379	
4.3	ASU/Oxidant Compression	\$199,700	\$0	\$0	\$0	\$0	\$199,700	\$19,357	\$0	\$21,906	\$240,962	\$522	
4.4	LT Heat Recovery & FG Saturation	\$27,678	\$0	\$26,458	\$0	\$0	\$54,135	\$5,284	\$0	\$11,883	\$71,302	\$154	
4.6	Flare Stack System	\$0	\$1,620	\$655	\$0	\$0	\$2,275	\$219	\$0	\$498	\$2,991	\$6	
4.9	Gasification Foundations	\$0	\$10,067	\$6,004	\$0	\$0	\$16,071	\$1,471	\$0	\$4,386	\$21,928	\$47	
	SUBTOTAL 4.	\$265,098	\$30,547	\$90,069	\$0	\$0	\$385,715	\$36,473	\$28,384	\$61,482	\$512,053	\$1,109	
5A	GAS CLEANUP & PIPING												
5A.1	Double Stage Selexol™	\$80,846	\$0	\$68,132	\$0	\$0	\$148,978	\$14,408	\$29,796	\$38,636	\$231,818	\$502	
5A.2	Elemental Sulfur Plant	\$4,987	\$972	\$6,391	\$0	\$0	\$12,350	\$1,200	\$0	\$2,710	\$16,260	\$35	
5A.3	Mercury Removal	\$3,273	\$0	\$2,474	\$0	\$0	\$5,747	\$555	\$287	\$1,317	\$7,905	\$17	
5A.4	Shift Reactors	\$8,901	\$0	\$3,559	\$0	\$0	\$12,460	\$1,195	\$0	\$2,731	\$16,386	\$35	
5A.6	Blowback Gas Systems	\$2,586	\$435	\$245	\$0	\$0	\$3,266	\$310	\$0	\$715	\$4,291	\$9	
5A.7	Fuel Gas Piping	\$0	\$939	\$614	\$0	\$0	\$1,554	\$144	\$0	\$340	\$2,038	\$4	
5A.9	HGCU Foundations	\$0	\$915	\$616	\$0	\$0	\$1,531	\$141	\$0	\$501	\$2,173	\$5	
	SUBTOTAL 5.	\$100,594	\$3,261	\$82,031	\$0	\$0	\$185,886	\$17,952	\$30,083	\$46,951	\$280,871	\$608	
5B	CO2 REMOVAL & COMPRESSION												
5B.1	CO2 Removal System	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
5B.2	CO2 Compression & Drying	\$36,579	\$0	\$12,219	\$0	\$0	\$48,798	\$4,698	\$0	\$10,699	\$64,195	\$139	
	SUBTOTAL 5B.	\$36,579	\$0	\$12,219	\$0	\$0	\$48,798	\$4,698	\$0	\$10,699	\$64,195	\$139	
6	COMBUSTION TURBINE/ACCESSORIES	\$111,211	\$923	\$8,948	\$0	\$0	\$121,083	\$11,476	\$11,909	\$14,882	\$159,350	\$345	
7	HRSg, DUCTING & STACK	\$32,150	\$2,972	\$9,341	\$0	\$0	\$44,462	\$4,202	\$0	\$5,623	\$54,287	\$118	
8	STEAM TURBINE GENERATOR	\$73,602	\$938	\$18,882	\$0	\$0	\$93,422	\$8,932	\$0	\$17,286	\$119,640	\$259	
9	COOLING WATER SYSTEM	\$5,621	\$7,506	\$6,893	\$0	\$0	\$20,020	\$1,861	\$0	\$4,692	\$26,573	\$58	
10	ASH/SPENT SORBENT HANDLING SYS	\$22,268	\$1,715	\$10,966	\$0	\$0	\$34,949	\$3,354	\$0	\$4,185	\$42,488	\$92	
11	ACCESSORY ELECTRIC PLANT	\$35,159	\$15,468	\$28,170	\$0	\$0	\$78,797	\$6,779	\$0	\$16,381	\$101,957	\$221	
12	INSTRUMENTATION & CONTROL	\$12,763	\$2,587	\$8,391	\$0	\$0	\$23,741	\$2,150	\$1,186	\$4,543	\$31,620	\$68	
13	IMPROVEMENTS TO SITE	\$3,698	\$2,179	\$9,703	\$0	\$0	\$15,580	\$1,539	\$0	\$5,135	\$22,254	\$48	
14	BUILDINGS & STRUCTURES	\$0	\$7,597	\$8,630	\$0	\$0	\$16,227	\$1,476	\$0	\$2,911	\$20,614	\$45	
CALCULATED TOTAL COST		\$841,832	\$96,963	\$369,906	\$0	\$0	\$1,308,701	\$122,057	\$89,428	\$251,446	\$1,771,631	\$3,837	

Table 5-8
Case 2c Total Plant Cost Summary

Owner's Costs	\$ x \$1,000	\$/kW
Preproduction Costs		
6 months All Labor	\$15,241	\$33
1 Month Maintenance Materials	\$3,318	\$7
1 Month Non-Fuel Consumables	\$466	\$1
1 Month Waste Disposal	\$425	\$1
25% of 1 Months Fuel Cost at 100% CF	\$1,008	\$2
2% of TPC	\$35,433	\$77
Total	\$55,890	\$121
Inventory Capital		
60 day supply of fuel at 100% CF	\$7,953	\$17
60 day supply of non-fuel consumables at 100% CF	\$690	\$1
0.5% of TPC (spare parts)	\$8,858	\$19
Total	\$17,501	\$38
Initial Cost for Catalyst and Chemicals	\$14,927	\$32
Land	\$900	\$2
Other Owner's Cost	\$265,745	\$575
Financing Costs	\$47,834	\$104
Total Owner's Costs	\$402,797	\$872
Total Overnight Costs (TOC)	\$2,174,429	\$4,709

5.7 OPERATING COSTS

Table 5-9 shows the operating cost breakdown for the Case 2c IGCC.

Table 5-9
Case 2c Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Case 2c - AR Compact Gasifier IGCC w/Selexol™-based AGR and CO2 Capture				
Plant Size (MWe)	462			Heat Rate (Btu/kWh):	10,436
Primary/Secondary Fuel:	PRB			Fuel Cost (\$/MMBtu):	
Design/Construction	5 years			Book Life (yrs):	20
TPC (Plant Cost) Year	June 2011			TPI Year:	2016
Capacity Factor (%)	80			CO2 Captured (TPD)	11137
<u>OPERATING & MAINTENANCE LABOR</u>					
Operating Labor					
Operating Labor Rate (base):	\$39.70	\$/hr			
Operating Labor Burden:	30.00	% of base			
Labor Overhead Charge	25.00	% of labor			
 Operating Labor Requirements per Shift	units/mod		Total Plant		
Skilled Operator	2.0		2.0		
Operator	10.0		10.0		
Foreman	1.0		1.0		
Lab Tech's etc	3.0		3.0		
TOTAL Operating Jobs	16.0		16.0		
				<u>Annual Cost</u>	<u>Annual Unit Cost</u>
				\$	\$/kW-net
Annual Operating Labor Cost				\$7,233,658	
Maintenance Labor Cost				\$17,151,162	
Administration & Support Labor				\$6,096,205	
Property Taxes and Insurance				\$35,432,624	
TOTAL FIXED OPERATING COSTS				\$65,913,648	
<u>VARIABLE OPERATING COSTS</u>					
Maintenance Material Cost				\$31,852,158	<u>\$/kWh-net</u>
<u>Consumables</u>	<u>Consumption</u>	<u>Unit</u>	<u>Initial Fill</u>		
	<u>Initial</u>	<u>/Day</u>	<u>Cost</u>	<u>Cost</u>	
Water(/1000 gallons)	0	2,532	1.67	\$0	\$1,237,737
Chemicals					
MU & WT Chem (lb)	0	15087	0.27	\$0	\$1,179,953
Carbon (Hg Removal) (lb)	111993	154	1.63	\$182,549	\$73,128
COS Catalyst (m3)	0	0	3751.70	\$0	\$0
Water Gas Shift Catalyst (ft3)	5981	4.09	771.99	\$4,617,230	\$922,803
Selexol™ Solution (gal)	275283	87.70	36.79	\$10,127,659	\$942,137
SCR Catalyst (m3)	0	0	0.00	\$0	\$0
Ammonia (19% NH3) (ton)	0	0	0.00	\$0	\$0
Claus Catalyst (ft3)	w/equip	1.92	203.15	\$0	\$113,988
Subtotal Chemicals				\$14,927,438	\$3,232,010
Other					
Supplemental Fuel (MMBtu)	0	0	0.00	\$0	\$0
Gases, N2 etc.(/100scf)	0	0	0.00	\$0	\$0
LP Steam (/1000 lbs)	0	0	0.00	\$0	\$0
Subtotal Other				\$0	\$0
Waste Disposal:					
Spent Mercury Catalyst (lb)	0	154	0.65	\$0	\$29,161
Flyash (ton)	0	0	0.00	\$0	\$0
Slag (ton)	0	553	25.11	\$0	\$4,053,550
Subtotal Waste Disposal				\$0	\$4,082,711
By-products & Emissions					
Sulfur (tons)	0	49	0.00	\$0	\$0
Subtotal By-Products				\$0	\$0
TOTAL VARIABLE OPERATING COSTS				\$14,927,438	\$40,404,616
Fuel (tons)	0	6752	19.63	\$0	\$38,703,730

5.8 COST OF ELECTRICITY

Table 5-10 shows a summary of the power output, CAPEX, OPEX, COE and cost of CO₂ capture for the Case 2c AR Compact Gasifier with Selexol™ AGR IGCC. The Case 2c IGCC COE is 128.3 mills/kWh

Table 5-10
Plant Performance and Economic Summary

Case	Case 2c
CAPEX, \$MM	
Total Installed Cost (TIC)	\$1,309
Total Plant Cost (TPC)	\$1,772
Total Overnight Cost (TOC)	\$2,174
OPEX, \$MM/yr (100% Capacity Factor Basis)	
Fixed Operating Cost (OC _{fix})	\$65.9
Variable Operating Cost, less Fuel (OC _{var})	\$50.5
Fuel (OC _{fuel})	\$48.4
Total OPEX	\$164.8
Power Production, MWe	
Gas Turbine	430.2
Steam Turbine	209.3
Auxiliary Power Consumption	177.8
Net Power Output	461.8
Power Generated, MWh/yr (MWh)	4,045,045
SO ₂ Emissions (lb/MWh _{gross})	0.0053
SO ₂ Emissions (lb/MMBtu)	0.0007
COE, excl CO₂ TS&M, mills/kWh	128.3
COE, incl CO₂ TS&M, mills/kWh	148.4
Cost of CO₂ Avoided excl CO₂ TS&M, \$/ton CO₂	59.0
Cost of CO₂ Avoided incl CO₂ TS&M, \$/ton CO₂	83.1

6.1 PROCESS OVERVIEW

The final case in this techno-economic evaluation, Case 2d Shell Gasifier with RTI WDP IGCC, has the same configuration as Case 2b, with the exception that the AR compact gasifier is replaced by the original Shell gasifier from Case 2a. It is designed to capture CO₂ equivalent to 90% of the raw syngas' carbon content using RTI's WDP in combination with the AACRP. The IGCC plant operates as a base-loaded unit with an annual on-stream CF of 80%.

Case 2d, the standalone RTI WDP/AACRP case, can be compared with Case 2b to identify any synergistic advantages in integrating both the AR and RTI systems in an IGCC plant.

The Case 2d IGCC has the following characteristics that differentiate it from the Case 2a Shell Reference IGCC:

- Syngas leaving the gasifier that has been quenched and filtered enters RTI's WDP for sulfur removal at about 760°F instead of going through a low temperature scrubber as per Case 2a.
- After sulfur removal in the WDP, the treated syngas in Case 2d then enters the WGS reactors. Unlike the Case 2a sour shift reactors, the Case 2d AFWGS consists of fixed-bed reactors (using commercial high-temperature sweet WGS catalyst) combined in such a manner as to significantly reduce the overall steam requirement and reactor capital cost while still meeting the catalyst manufacturer's steam to CO recommendations. These reactors operate at standard temperatures for commercial high-temperature sweet WGS processes, but at a higher inlet temperature than commercial sour WGS processes. Commercial sweet WGS catalyst cost about half as much as commercial sour WGS catalysts.
- After the hydrogen-rich shifted syngas is cooled, it enters the AACRP for CO₂ capture. Unlike Selexol™, which is a physical solvent, activated methyldiethanolamine is a chemical solvent. As ~99.9% of the sulfur compounds have been removed by the WDP process upstream already, the amine process only has to remove CO₂ and is less complicated than the two-stage Selexol™ setup. However, the AACRP unit does also capture ~99% of any residual sulfur left in the syngas following WDP along with the CO₂, resulting in an overall system total sulfur reduction of >99.99% (sub-ppmv total sulfur in the final cleaned syngas).
- The DSRP replaces the Claus process in Case 2a. In the DSRP, sulfur leaving the WDP process in the form of SO₂ is reduced by a slipstream of shifted, hydrogen-rich syngas, forming elemental sulfur, H₂S and COS. The elemental sulfur is condensed while the remaining H₂S, and COS are re-oxidized in the presence of air to SO₂. The SO₂ is then removed in a lime scrubber, forming gypsum (CaSO₄·2H₂O).

The coal feed rate is expected to be the same as Case 2a since Case 2d also utilizes the Shell gasifier operating under the same conditions. It therefore has the same cold gas efficiency and generates the same amount of raw gasifier syngas that is eventually fired in the two GE 7F-turbines.

6.2 CASE 2d PROCESS DESCRIPTION

The system description below for Case 2d follows the BFD in Figure 6-1 and stream numbers reference the same figure. Table 6-1 provides the Nexant generated process data for the numbered streams in the BFD.

Coal Preparation and Drying

Same as Case 2a

Coal Feed System

Same as Case 2a

Air Separation Unit

Same as Case 2b

Shell Gasifier

Same as Case 2a, except no raw gas cooling. The syngas after quench is routed directly to RTI's WDP section at 760°F.

Warm Syngas Cleanup Process

The Case 2d RTI advanced syngas cleanup process consists of the following major system components: WDP, DSRP, AFWGS, and LTGC. AACRP is integrated with the RTI advanced syngas cleanup process for CO₂ removal.

Warm Gas Desulfurization Process

Same as Case 2b

Direct Sulfur Recovery Process

Same as Case 2b

Advance Fixed-Bed Water Gas Shift

Same as Case 2b

Low Temperature Gas Cooling and Mercury Removal

Same as Case 2b

Activated Amine CO₂ Removal Process

Same as Case 2b

CO₂ Compression and Dehydration

Same as Case 2a

Power Block

The treated syngas leaving the AACRP is reheated to 385°F. It is then diluted with nitrogen and SO₂ scrubber overhead gas from the DSRP to meet the heating value specifications before it enters the GT combustor. The exhaust gas exits the GT at a nominal 1,050°F and enters the HRSG where additional heat is recovered. The flue gas exits the HRSG at 270°F and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced commercially available steam turbine using a nominal 1,800 psig/1,000°F/1,000°F steam cycle. There is no air integration between the gas turbine and the ASU.

Balance of Plant

Balance of plant items are covered in Sections 3.2.9 through 3.2.12.

Figure 6-1
Case 2d: Shell with RTI WDP IGCC Simplified BFD

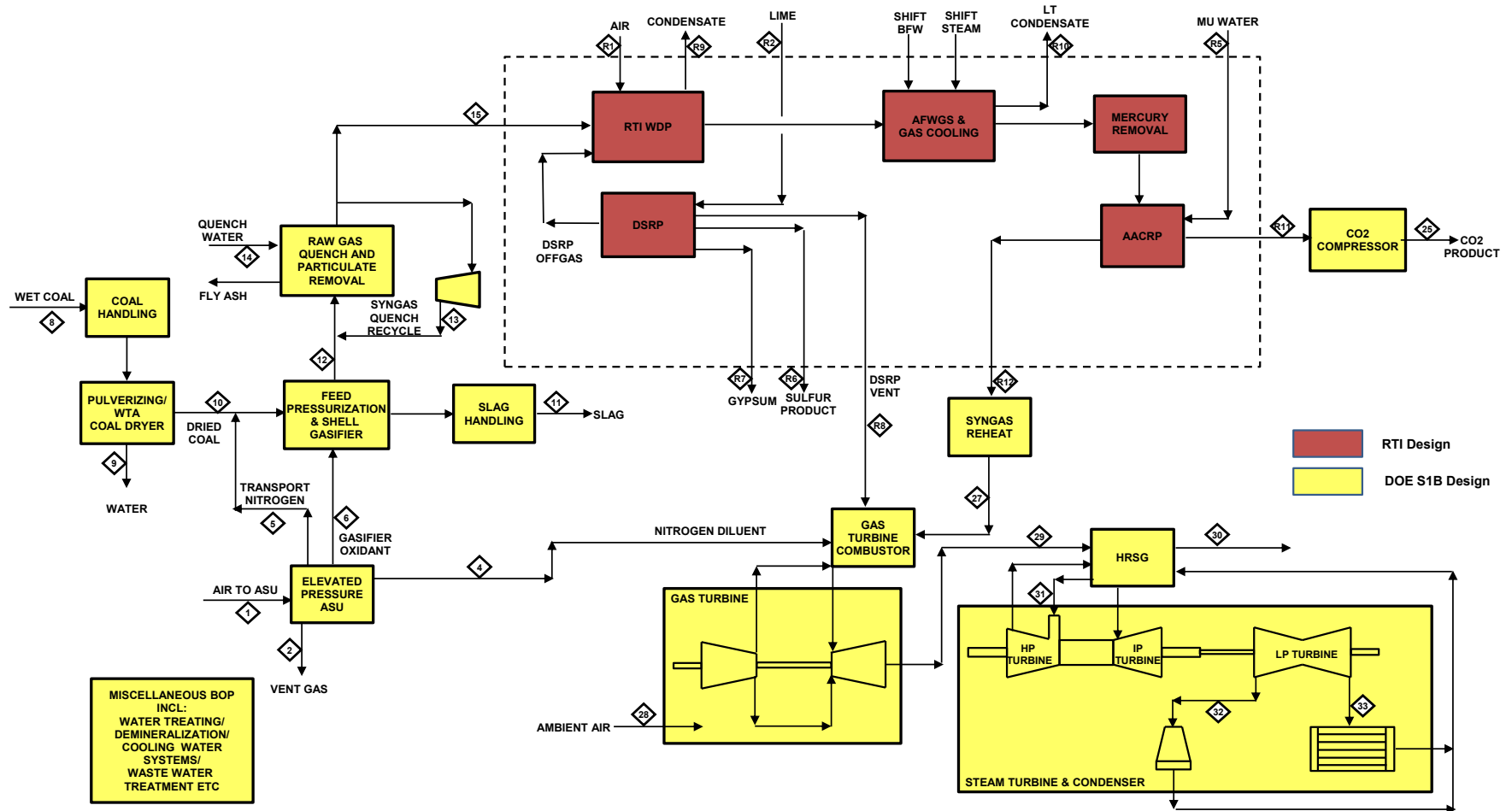


Table 6-1
Case 2d Stream Table

Description	Air to ASU	ASU Vent Gas	Nitrogen Diluent	Transport Nitrogen	Gasifier Oxidant	Wet Coal	Coal Moisture	Dried Coal	Slag	Hot Syngas	Syngas Recycle	Quench Water	Cooled Syngas
Stream No.	1	2	4	5	6	8	9	10	11	12	13	14	15
V-L Mole Fraction													
Ar	0.0093	0.0165	0.0023	0.0023	0.0318	0	0	0	0	0.0093	0.0054	0	0.0054
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0	0.6010	0.3466	0	0.3466
CO2	0.0003	0.0047	0	0	0	0	0	0	0	0.0327	0.0188	0	0.0188
COS	0	0	0	0	0	0	0	0	0	0.0005	0.0003	0	0.0003
H2	0	0	0	0	0	0	0	0	0	0.2609	0.1504	0	0.1504
H2O	0.0064	0.0988	0.000	0	0	0	1.0000	0	0	0.0287	0.4392	0.9981	0.4392
H2S	0	0	0	0	0	0	0	0	0	0.0030	0.0017	0	0.0017
N2	0.7759	0.7586	0.9921	0.9921	0.0178	0	0	0	0	0.0609	0.0351	0	0.0351
NH3	0	0	0	0	0	0	0	0	0	0.0029	0.0025	0.0019	0.0025
O2	0.2081	0.1213	0.0054	0.0054	0.9504	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0
CaSO4	0	0	0	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.000	1.0000	1.0000	1.0000	0	1.0000	0	0	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	53470	3384	36971	2061	11054	0	6841	0	0	38321	18310	28138	66459
V-L Flowrate, lb/hr	1545000	94029	1037410	57841	355725	0	123241	0	0	826852	367452	506861	1333710
Solids Flow Rate, lb/hr	0	0	0	0	0	585971	0	462730	49444	0	0	0	0
Temperature, F	42	64	385	387	292		75			2602	781	348	765
Pressure, psia	13.0	16.4	384.0	815.0	740.0	13.0	48.0	14.8	615.0	615.0	615.0	600.0	590.0
Enthalpy, Btu/lb	-33.3	-463.2	75.1	74.8	43.8		-6823			-798	-2993	-6538	-3000
Density, lb/cuft	0.070	0.089	1.180	2.473	2.961		62.273			0.401	0.930	55.788	0.905

Case 2d Stream Table (cont'd)

Description	WGD Air	Lime	MDEA MU Water	Sulfur Product	Gypsum	DSRP Vent	WDP Cond	Syngas Cool Cond	MDEA CO2 Out	Treated Syngas
Stream No.	R1	R2	R5	R6	R7	R8	R9	R10	R11	R12
V-L Mole Fraction										
Ar	0.0093	0	0	0	0	0.0134	0	0	0.0000	0.0100
CH4	0	0	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0.0001	0.0655
CO2	0.0003	0	0	0	0.0035	0.0197	0	0.0002	0.9392	0.0012
COS	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0.0014	0.8504
H2O	0.0064	0.9737	1.0000	0	0.9769	0.0005	1.0000	0.9929	0.0593	0.0041
H2S	0	0	0	0	0	0	0	0	0	0.0000
N2	0.7759	0	0	0	0	0.9661	0	0	0.0000	0.0687
NH3	0	0	0	0	0	0	0	0.0070	0	0
O2	0.2081	0	0	0	0	0.0003	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0
S	0	0	0	1.0000	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0.0263	0	0	0.0013	0	0	0	0	0
CaSO4	0	0	0	0	0.0180	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	1184	904	1428	109	1113	827	3	58154	23315	35712
V-L Flowrate, lb/hr	34208	17626	25732	3486	22600	23559	58	1184120	988795	214335
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0
Temperature, F	42	75	122	458	99	103	80	95	122	122
Pressure, psia	13.0	13.0	471.0	447.0	15	470.0	13.0	480.6	30.7	471.0
Enthalpy, Btu/lb	-33		-6817	2620		-116	-6863	-3251	-3887	-575
Density, lb/cuft	0.070		60.524	118.692		2.235	61.603	1.676	0.212	0.447

Case 2d Stream Table (cont'd)

Description	CO2 Product	Reheat Syngas	Ambient Air to GT	GT Flue Gas	Stack Gas	HP SH Steam	Air-Cooled LP Cond	Water-Cooled LP Cond	BFW to HRSG
Stream No.	25	27	28	29	30	31	32	33	34
V-L Mole Fraction									
Ar	0	0.0100	0.0093	0.0090	0.0090	0.0000	0	0	0
CH4	0	0	0	0	0	0	0	0	0
CO	0.0001	0.0655	0.0000	0	0	0	0	0	0
CO2	0.9984	0.0012	0.0003	0.0089	0.0089	0.0000	0	0	0
COS	0	0	0	0	0	0	0	0	0
H2	0.0014	0.8504	0.0000	0	0	0	0	0	0
H2O	0.0000	0	0.0064	0.1147	0.1147	1.0000	1.0000	1.0000	1.0000
H2S	0	0	0	0	0	0	0	0	0
N2	0.0000	0.0687	0.7759	0.7600	0.7600	0.0000	0	0	0
NH3	0	0	0	0	0	0	0	0	0
O2	0	0	0.2081	0.1073	0.1073	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0
CaSO4	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	21933	35712	221242	278397	278397	54806	46424	46424	91421
V-L Flowrate, lb/hr	963904	214335	6392720	7668020	7668020	987347	836350	836350	1646970
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0
Temperature, F	162	385	42	1057	270	999	90	90	241
Pressure, psia	2214.7	461.0	12.9	13.5	13.0	1815.0	0.7	0.7	105.0
Enthalpy, Btu/lb		-266	-33	-224	-438	-5387	-6808	-6808	-6657
Density, lb/cuft	28.796	0.302	0.069	0.023	0.046	2.282	31.058	31.058	59.075

6.3 SPARING PHILOSOPHY

The sparing philosophy for Case 2d is provided below. Single train designs were utilized throughout with exceptions where equipment had exceeded its maximum capacity limit or dimensions for acceptable transportation to the site and thus required an additional train. For AACRP, multiple trains were used for specific groups of equipment to enable meeting the transportation dimension constraints. The design has:

- Two ASU trains (2 x 50%)
- Two trains of coal drying (2 x 50%)
- Two trains of gasification, including dry feed system, Shell gasifier, syngas quench cooler (2 x 50%)
- One WDP train (1 x 100%)
- One DSRP train (1 x 100%)
- One AFWGS and LTGC train (1 x 100%)
- One AACRP train (3 x 33% for rich amine absorber and LP flash drum, 1 x 100% for rest of process)
- Two gas turbines/HRSG tandems (2 x 50%)
- One steam turbine (1 x 100%)

6.4 PERFORMANCE RESULTS

The Nexant-modeled Case 2d IGCC plant with CO₂ capture consumes 7,032 tpd of PRB coal at the Montana site to produce a net output of 464 MWe at a net plant efficiency of 31.53% on a HHV basis. Overall performance for the Case 2d IGCC plant is summarized in Table 6-2, which includes auxiliary power requirements. Loads that are unique to the RTI advanced syngas cleanup and AACRP systems are shown in bold and italicized.

As in Case 2b, essentially no effort was made to optimize the pressure drop profile for the entire system to compensate for the removal of the heating and cooling equipment to optimize the process configuration for integration of the RTI WDP for Case 2d. In this case, the integration with the Shell gasifier resulted in similar operating pressures for Selexol™ in the reference case (Case 2a) and AACRP in this case (Case 2d). The improvement in the net efficiency percentage between these two cases was about 0.21. From a comparison of Cases 2a and 2d and Cases 2b and 2c, RTI's warm syngas cleanup process increases net efficiency by between 0.21 and 0.12 percentage points, respectively. These results demonstrate that the net efficiency of the integration of both advanced technologies due to efficiency improvements from both technologies.

Table 6-2
Case 2d Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case 2d
Gas Turbine Power	429,570
Steam Turbine Power	226,366
TOTAL POWER, kWe	665,936
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	510
Coal Milling	2,730
Slag Handling	580
WTA Coal Dryer Compressor	9,370
WTA Coal Dryer Auxiliaries	620
Air Separation Unit Auxiliaries	985
Air Separation Unit Main Air Compressor	62,576
Oxygen Compressor	8,830
Nitrogen Compressors	32,725
CO ₂ Compressor	33,103
Boiler Feedwater Pumps	4,252
Condensate Pump	230
Quench Water Pump	760
Syngas Recycle Compressor	1,019
Circulating Water Pump	3,026
Ground Water Pumps	335
Cooling Tower Fans	1,973
Air Cooled Condenser Fans	2,927
AACRP	16,407
Gas Turbine Auxiliaries	997
Steam Turbine Auxiliaries	97
RTI WDP	2,311
DSRP	283
Miscellaneous Balance of Plant	3,000
Transformer Losses	2,521
TOTAL AUXILIARIES, kWe	192,167
NET POWER, kWe	463,769
Net Plant Efficiency, % (HHV)	31.53%
Net Plant Heat Rate, Btu/kWh	10,821
CONDENSER COOLING DUTY, MMBtu/hr	1,236
CONSUMABLES	
As-Received Coal Feed, lb/hr	585,971
Thermal Input, kWt	1,470,705
Raw Water Withdrawal, gpm	3,674
Raw Water Consumption, gpm	2,975

Table 6-3 shows the carbon balance for the Case 2d IGCC plant. The carbon capture efficiency is defined as the amount of carbon in the CO₂ product stream relative to the amount of carbon in the coal less carbon contained in the slag. For Case 2d, the carbon capture efficiency is 90.1%.

Table 6-3
Case 2d Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	293,384	
ASU Air	193	
Air to WDP	4	
Air to Gas Turbine	797	
ASU Vent		193
LTGC Condensate to Cooling Tower		45
Carbon in Slag		1,467
Stack Gas		29,635
CO ₂ Product		263,038
Convergence Tolerance		0
Total	294,378	294,378

Table 6-4 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the elemental sulfur condensed in the DSRP, gypsum from the lime scrubber, sulfur emitted in the stack gas, and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible.

Table 6-4
Case 2d Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	4,263	
Gypsum		764
Sulfur Product		3,486
Stack Gas		0
CO ₂ Product		13
Convergence Tolerance		0
Total	4,263	4,263

Table 4-5 shows the overall water balance for the Case 2d IGCC plant. The raw water withdrawal is the net difference between the water demand and the internal recycle. The net water consumption is the difference between the raw water withdrawal and process water discharge.

Table 6-5
Case 2d Overall Water Balance

Water Use, gpm	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Net Water Consumption
Slag Handling	128	128	0	0	0
Quench/Wash	1,013	765	248	0	248
KO Drum Blowdown	0	0	0	0	0
Condenser Makeup	599	0	599	0	599
Shift Steam	356	0	356		
Shift BFW	180	0	180		
BFW Makeup	12	0	12		
AACRP Scrubber Makeup	51	0	51		
Cooling Tower	3,085	258	2,827	699	2,128
Water from Coal Drying		246	-246		
BFW Blowdown		12	-12		
Total	4,825	1,151	3,674	699	2975

6.5 EQUIPMENT LIST

The equipment list for the RTI processes that differ from the reference Shell Gasifier with Selexol™ AGR IGCC case is shown in Table 6-6 below.

Table 6-6
Case 2d Equipment List

ACCOUNT 5A **SYNGAS CLEANUP**
Subaccount 5A.1 **RTI WDP**

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	WDP Adsorption Reactor	Proprietary Transport Reactor	27,400 acfm	1	0
2	WDP Regeneration Reactor	Proprietary Transport Reactor	680 acfm	1	0
3	Adsorption Cyclone Separator	Cyclone Separator	27,400 acfm	1	0
4	Stripper Cyclone Separator	Cyclone Separator	680 acfm	1	0
5	Adsorber Filter	Candle Filter	27,400 acfm	1	0
6	Regenerator Offgas Filter		680 acfm	1	0
7	Adsorber Filter Lock Hopper + Fines Bin		Proprietary	1	0
8	Regenerator Filter Lock Hopper + Fines Bin		Proprietary	1	0
9	Sorbent Feeder Package (incl Hopper)		Proprietary	1	0
10	HP Steam Generator	Shell and Tube	84,100 lb/hr HP Steam	1	0
11	Regenerator Air Heat Exchanger	Shell and Tube	27,600 lb/hr	1	0
12	Recycle Syngas Compressor	Centrifugal	45 acfm @ 475 psia	1	0
13	Regenerator Air Compressor	Centrifugal	8,165 acfm @ 13 psia	1	0
14	DSRP Offgas Compressor	Centrifugal	95 acfm @ 470 psia	1	0
15	Syngas Recycle Compressor	Centrifugal	5 acfm @ 560 psia	1	0

ACCOUNT 5A SYNGAS CLEANUP
Subaccount 5A.2 RTI DSRP

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	DSRP Fixed Bed Reactor	Packed Bed Reactor	123,000 lb/hr	1	0
2	Tail Gas Oxidation Reactor	Packed Bed Reactor	123,000 lb/hr	1	0
3	DSRP Feed Gas Preheater	Shell and Tube	4.8 MMBtu/hr	1	0
4	Sulfur Condenser	Shell and Tube	3,500 lb/hr sulfur	1	0
5	Oxidation Reactor Gas Preheater	Shell and Tube	2.8MMBtu/hr	1	0
6	Liquid Sulfur Separator	Pressure Vessel	3,500 lb/hr sulfur	1	0
7	Recycle Gas Compressor	Centrifugal	850 acfm @ 407 psia	1	0
8	Pulsing Gas Compressor	Centrifugal	Average 11 acfm @ 470 psia	1	0
9	Scrubber Recycle Cooler	Shell and Tube	18.2 MMBtu/hr	1	0
10	SO ₂ Scrubber	Tray Column	1,520 lb/hr SO ₂	1	0
11	Recycle Pump	Centrifugal	1,500 gpm @ 150 ft H ₂ O	1	0
12	Lime Slurry Pump	Centrifugal	50 gpm @ 950 ft H ₂ O	1	0
13	Gypsum Filter	Bag Filter	1,200 acfm	1	0
14	Lime Makeup Drum	Pressure Vessel	16,000 lb/hr	1	0

ACCOUNT 5A SYNGAS CLEANUP
Subaccount 5A.4 Shift Reactors

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	WGS Reactor 1	Packed Bed Reactor	Proprietary	1	0
2	WGS Reactor 2	Packed Bed Reactor	Proprietary	1	0
3	WGS Reactor 3	Packed Bed Reactor	Proprietary	1	0

ACCOUNT 5B CO₂ REMOVAL & COMPRESSION
Subaccount 5B.1 AACRP CO₂ Removal System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	Rich Amine Absorber	Packed Column	241,800 cuft/hr syngas	3	0
2	Lean Amine Absorber	Packed Column w/ Wash Trays	559,800 cuft/hr syngas	1	0
3	Amine Stripper	Packed Column	1,749,600 cuft/hr gas flow	1	0
4	LP Flash	Packed Column	1,836,400 cuft/hr gas flow	3	0
5	LP Flash Reflux Drum	Reflux Drum	1,564,700 cuft/hr CO ₂	3	0
6	Lean/Rich Amine Exchanger	Shell and Tube	573,500 lb/hr	6	0
7	Amine Stripper Reboiler	Shell and Tube	563,500 lb/hr amine	6	0
8	Lean Amine Cooler	Shell and Tube	1,746,700 lb/hr amine	2	0
9	Flash Overhead Condenser	Shell and Tube	354,600 lb/hr CO ₂	3	0
10	MP Flash Condenser	Shell and Tube	7,300 lb/hr amine	1	0
11	Lean Amine Pump	Centrifugal	2,060 gpm	3	0
12	Semi-lean Amine Pump	Centrifugal	14,030 gpm	3	1
13	Rich Amine Pump	Centrifugal	10,070	1	0
14	Flash Reflux Pump	Centrifugal	51 gpm	3	0
15	Amine Stripper Bottoms Pump	Centrifugal	6,035 gpm	1	1
16	Filter Pump	Centrifugal	670 gpm	1	0
17	MP Flash Recycle KO Drum	KO Drum	22,260 cuft/hr	1	0
18	MP Flash Drum	Reflux Drum	131,900 cuft/hr	1	0
19	Recycle Gas Compressor	Centrifugal	371 acfm @ 160 psia	1	0
20	Amine Tank	Vertical Vessel	14,700 gallons	1	0
21	Filter Package	Filter Package	670 gpm	1	1
22	Filter Package Vessel	Vessel	3,200 gallons	1	1

6.6 CAPITAL COST

Table 6-7 shows the cost breakdown of the Case 2d Shell Gasifier with RTI WDP IGCC, expressed in a consistent format with the code of accounts in the NETL Report 1399. The accounts/subaccounts of interest for this study are:

- 2.3 Dry coal injection system,
- 4 Gasifier & accessories,
- 5A Gas cleanup & piping, and
- 5B CO₂ removal and compression.

These are shown with more detail to include the various subaccounts and provide more clarity to the major cost differences among the cases.

Table 6-8 shows the calculation and addition of owner's costs to determine the TOC, used to calculate COE.

The estimated TOC, in 2011 dollars, of the Case 2d Shell Gasifier with RTI WDP IGCC using PRB coal is \$5,054/kW.

Table 6-7
Case 2d Total Plant Cost Summary

Case 2d: Shell IGCC with RTI WDP Low Rank Western Coal Baseline Study		PRB	Coal Feed, lb/hr Coal HHV, Btu/lb		585,971 8564	Plant Size Net Efficiency		463.8 MW, net 31.5%			
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost \$	Eng'g CM H.O & Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect			Process	Project	\$	\$/kW
	1 COAL & SORBENT HANDLING	\$19,442	\$3,409	\$14,872	\$0	\$37,723	\$3,423	\$0	\$8,228	\$49,375	\$106
	2 COAL & SORBENT PREP & FEED										
	2.3 Dry Coal Injection System	\$57,741	\$0	\$39,047	\$0	\$96,788	\$8,337	\$0	\$21,025	\$126,150	\$272
	2.x Other Coal & Sorbent Prep & Feed Systems	\$62,434	\$10,880	\$17,044	\$0	\$90,358	\$7,894	\$0	\$19,649	\$117,901	\$254
	SUBTOTAL 2.	\$120,176	\$10,880	\$56,091	\$0	\$187,147	\$16,230	\$0	\$40,674	\$244,051	\$526
	3 FEEDWATER & MISC BOP SYSTEMS	\$9,596	\$7,461	\$9,364	\$0	\$26,421	\$2,493	\$0	\$6,672	\$35,587	\$77
	4 GASIFIER & ACCESSORIES										
	4.1 Gasifier, Syngas Cooler & Auxiliaries (Shell)	\$189,728	\$0	\$81,332	\$0	\$271,060	\$24,215	\$37,263	\$51,121	\$383,659	\$827
	4.3 ASU/Oxidant Compression	\$219,049	\$0	\$0	\$0	\$219,049	\$21,232	\$0	\$24,028	\$264,310	\$570
	4.4 LT Heat Recovery & FG Saturation	\$21,865	\$0	\$20,889	\$0	\$42,753	\$4,173	\$0	\$9,385	\$56,311	\$121
	4.6 Flare Stack System	\$0	\$1,721	\$696	\$0	\$2,417	\$232	\$0	\$529	\$3,179	\$7
	4.9 Gasification Foundations	\$0	\$10,698	\$6,380	\$0	\$17,078	\$1,564	\$0	\$4,660	\$23,302	\$50
	SUBTOTAL 4.	\$430,642	\$12,419	\$109,297	\$0	\$552,358	\$51,416	\$37,263	\$89,724	\$730,761	\$1,576
	5A GAS CLEANUP & PIPING										
	5A.1 RTI WDP	\$24,690	\$0	\$29,628	\$0	\$54,318	\$5,253	\$10,864	\$14,087	\$84,522	\$182
	5A.2 RTI DSRP	\$6,324	\$0	\$7,000	\$0	\$13,324	\$1,294	\$2,665	\$3,457	\$20,740	\$45
	5A.3 Mercury Removal	\$3,489	\$0	\$2,637	\$0	\$6,125	\$591	\$306	\$1,404	\$8,426	\$18
	5A.4 RTI Shift Reactors	\$3,214	\$0	\$1,286	\$0	\$4,500	\$431	\$0	\$986	\$5,917	\$13
	5A.6 Blowback Gas Systems	\$2,729	\$459	\$258	\$0	\$3,447	\$327	\$0	\$755	\$4,529	\$10
	5A.7 Fuel Gas Piping	\$0	\$950	\$621	\$0	\$1,571	\$146	\$0	\$344	\$2,060	\$4
	5A.9 Gas Clean Up Foundations	incl w/ WDP	\$0	incl w/ WDP	\$0	\$0	\$0	\$0	\$0	\$0	\$0
	SUBTOTAL 5.	\$40,446	\$1,409	\$41,430	\$0	\$83,285	\$8,043	\$13,834	\$21,032	\$126,195	\$272
	5B CO2 REMOVAL & COMPRESSION										
	5B.1 AACRP	\$25,067	\$0	\$21,056	\$0	\$46,123	\$4,461	\$9,225	\$11,962	\$71,770	\$155
	5B.2 CO2 Compression & Drying	\$39,403	\$0	\$13,162	\$0	\$52,566	\$5,061	\$0	\$11,525	\$69,151	\$149
	SUBTOTAL 5B.	\$64,470	\$0	\$34,218	\$0	\$98,689	\$9,521	\$9,225	\$23,486	\$140,921	\$304
	6 COMBUSTION TURBINE/ACCESSORIES	\$111,211	\$923	\$8,948	\$0	\$121,083	\$11,476	\$11,909	\$14,882	\$159,350	\$344
	7 HRSG, DUCTING & STACK	\$31,865	\$2,984	\$9,298	\$0	\$44,147	\$4,172	\$0	\$5,591	\$53,910	\$116
	8 STEAM TURBINE GENERATOR	\$77,732	\$992	\$19,834	\$0	\$98,559	\$9,433	\$0	\$18,186	\$126,179	\$272
	9 COOLING WATER SYSTEM	\$5,835	\$7,760	\$7,127	\$0	\$20,722	\$1,926	\$0	\$4,854	\$27,502	\$59
	10 ASH/SPENT SORBENT HANDLING SYS	\$23,279	\$1,783	\$11,463	\$0	\$36,525	\$3,505	\$0	\$4,372	\$44,402	\$96
	11 ACCESSORY ELECTRIC PLANT	\$36,075	\$16,008	\$29,065	\$0	\$81,148	\$6,983	\$0	\$16,896	\$105,027	\$226
	12 INSTRUMENTATION & CONTROL	\$12,897	\$2,614	\$8,479	\$0	\$23,991	\$2,173	\$1,199	\$4,591	\$31,954	\$69
	13 IMPROVEMENTS TO SITE	\$3,728	\$2,197	\$9,782	\$0	\$15,708	\$1,551	\$0	\$5,178	\$22,437	\$48
	14 BUILDINGS & STRUCTURES	\$0	\$7,691	\$8,737	\$0	\$16,428	\$1,495	\$0	\$2,945	\$20,868	\$45
CALCULATED TOTAL COST		\$987,396	\$78,531	\$378,007	\$0	\$1,443,934	\$133,842	\$73,430	\$267,313	\$1,918,518	\$4,137

Table 6-8
Case 2d Total Plant Cost Summary

Owner's Costs	\$ x \$1,000	\$/kW
Preproduction Costs		
6 months All Labor	\$16,129	\$35
1 Month Maintenance Materials	\$3,593	\$8
1 Month Non-Fuel Consumables	\$548	\$1
1 Month Waste Disposal	\$456	\$1
25% of 1 Months Fuel Cost at 100% CF	\$1,050	\$2
2% of TPC	\$38,370	\$83
Total	\$60,146	\$130
Inventory Capital		
60 day supply of fuel at 100% CF	\$8,282	\$18
60 day supply of non-fuel consumables at 100% CF	\$548	\$1
0.5% of TPC (spare parts)	\$9,593	\$21
Total	\$18,423	\$40
Initial Cost for Catalyst and Chemicals	\$6,152	\$13
Land	\$900	\$2
Other Owner's Cost	\$287,778	\$621
Financing Costs	\$51,800	\$112
Total Owner's Costs	\$425,199	\$917
Total Overnight Costs (TOC)	\$2,343,717	\$5,054

6.7 OPERATING COSTS

Table 6-9 shows the operating cost breakdown for the Case 2d IGCC.

Table 6-9
Case 2d Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Case 2d - Shell IGCC with RTI WDP				
Plant Size (MWe)	464	Heat Rate (Btu/kWh):		10,821	
Primary/Secondary Fuel:	PRB	Fuel Cost (\$/MMBtu):			
Design/Construction	5 years	Book Life (yrs):		20	
TPC (Plant Cost) Year	June 2011	TPI Year:		2016	
Capacity Factor (%)	80	CO2 Captured (TPD)		11565	
OPERATING & MAINTENANCE LABOR					
Operating Labor				TPC	\$1,918,518
Operating Labor Rate (base):		\$39.70	\$/hr		
Operating Labor Burden:		30.00 % of base			
Labor Overhead Charge		25.00 % of labor			
Operating Labor Requirements per Shift		units/mod	Total Plant		
Skilled Operator		2.0	2.0		
Operator		10.0	10.0		
Foreman		1.0	1.0		
Lab Tech's etc		3.0	3.0		
TOTAL Operating Jobs		16.0	16.0		
				Annual Cost	Annual Unit Cost
				\$	\$/kW-net
Annual Operating Labor Cost				\$7,233,658	
Maintenance Labor Cost				\$18,573,176	
Administration & Support Labor				\$6,451,708	
Property Taxes and Insurance				\$38,370,366	
TOTAL FIXED OPERATING COSTS				\$70,628,908	
VARIABLE OPERATING COSTS					
Maintenance Material Cost				\$34,493,041	\$/kWh-net
Consumables	Consumption		Unit	Initial Fill	
	Initial	/Day	Cost	Cost	
Water(/1000 gallons)	0	2,646	1.67	\$0	\$1,293,215
Chemicals					
MU & WT Chem (lb)	0	15763	0.27	\$0	\$1,232,841
Carbon (Hg Removal) (lb)	116627	160	1.63	\$190,102	\$76,154
RTI WDP Sorbent (lb)	Proprietary	Proprietary	Proprietary	\$2,582,019	\$1,618,436
RTI DSRP Catalyst (lb)	22000	12.1	11.15	\$245,300	\$39,248
DSRP Oxidation/Reduction Catalyst (lb)	22000	12.1	1.56	\$34,320	\$5,491
Lime	0	44448	0.0382	\$0	\$495,935
Water Gas Shift Catalyst (ft3)	5590	3.1	394.38	\$2,204,589	\$352,734
Activated Amine (lb)	320035	175	2.80	\$895,418	\$143,267
Subtotal Chemicals				\$6,151,748	\$3,964,107
Other					
Supplemental Fuel (MMBtu)	0	0	0.00	\$0	\$0
Gases, N2 etc. (/100scf)	0	0	0.00	\$0	\$0
LP Steam (/1000 lbs)	0	0	0.00	\$0	\$0
Subtotal Other				\$0	\$0
Waste Disposal:					
Spent Mercury Catalyst (lb)	0	160	0.65	\$0	\$30,368
Flyash (ton)	0	0	0.00	\$0	\$0
Slag (ton)	0	593	25.11	\$0	\$4,350,327
Subtotal Waste Disposal				\$0	\$4,380,695
By-products & Emissions					
Sulfur (tons)	0	42	0.00	\$0	\$0
Gypsum (tons)	0	271	0.00	\$0	\$0
Subtotal By-Products				\$0	\$0
TOTAL VARIABLE OPERATING COSTS				\$6,151,748	\$44,131,058
Fuel (tons)	0	7032	19.63	\$0	\$40,305,148

6.8 COST OF ELECTRICITY

Table 6-10 shows a summary of the power output, CAPEX, OPEX, COE and cost of CO₂ capture for the Case 2d Shell Gasifier with RTI WDP IGCC. The COE for the Case 2d IGCC is 137.3 mills/kWh

Table 6-10
Plant Performance and Economic Summary

Case	Case 2d
CAPEX, \$MM	
Total Installed Cost (TIC)	\$1,444
Total Plant Cost (TPC)	\$1,919
Total Overnight Cost (TOC)	\$2,344
OPEX, \$MM/yr (100% Capacity Factor Basis)	
Fixed Operating Cost (OC _{fix})	\$70.6
Variable Operating Cost, less Fuel (OC _{var})	\$55.2
Fuel (OC _{fuel})	\$50.4
Total OPEX	\$176.2
Power Production, MWe	
Gas Turbine	429.6
Steam Turbine	226.4
Auxiliary Power Consumption	192.2
Net Power Output	463.8
Power Generated, MWh/yr (MWh)	4,062,618
SO ₂ Emissions (lb/MWh _{gross})	0.0014
SO ₂ Emissions (lb/MMBtu)	0.0002
COE, excl CO₂ TS&M, mills/kWh	137.3
COE, incl CO₂ TS&M, mills/kWh	158.1
Cost of CO₂ Avoided excl CO₂ TS&M, \$/ton CO₂	70.0
Cost of CO₂ Avoided incl CO₂ TS&M, \$/ton CO₂	95.0

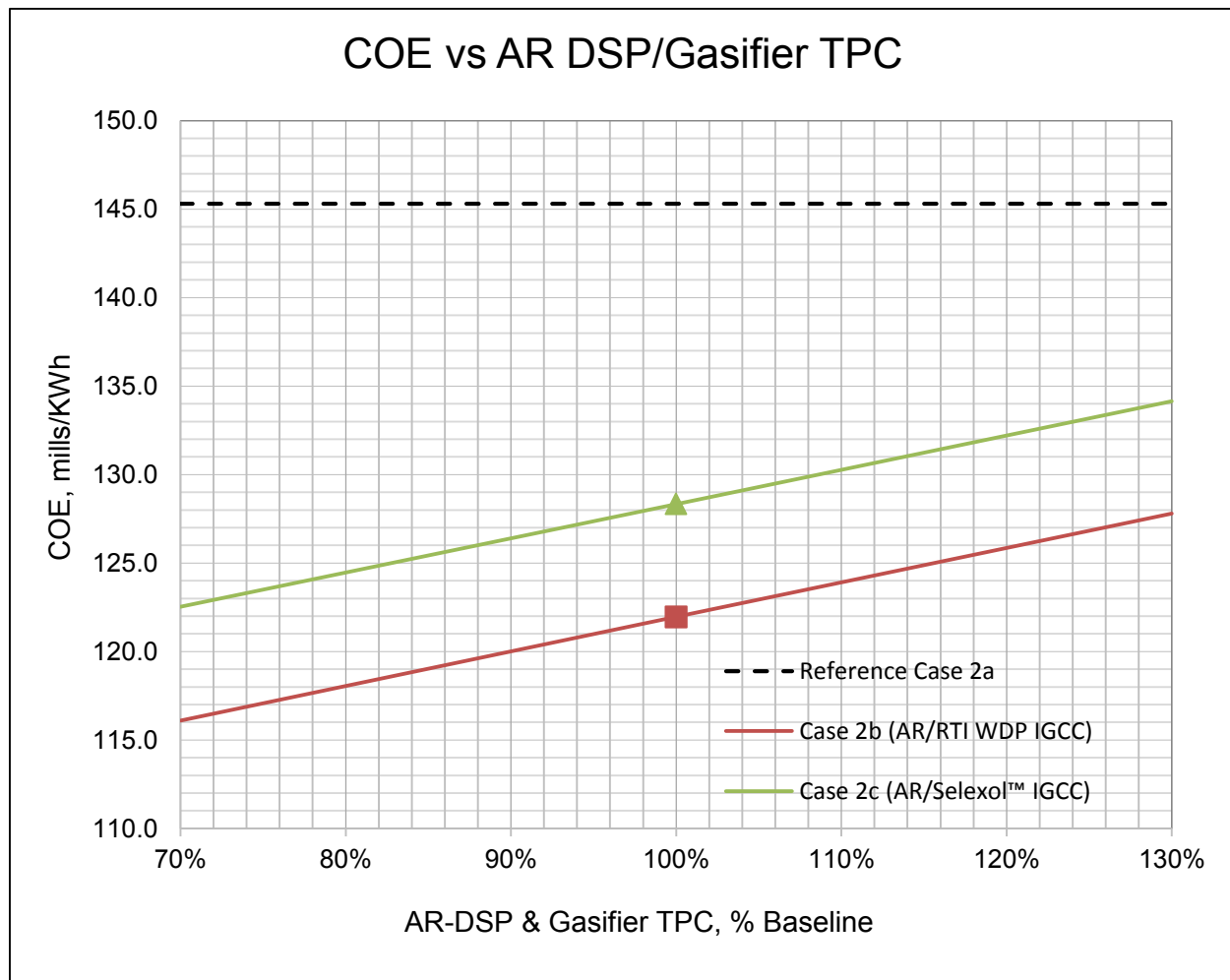
Section 7 Sensitivity Analysis

Sensitivity analysis was carried out to determine the effects of various parameters of the AR advanced gasification system and RTI advanced syngas cleanup and AACRP on the overall IGCC COE. The parameters investigated include: system capital cost, feedstock cost, IGCC plant capacity factor, CO₂ sales price, and cost of CO₂ emissions.

7.1 AR DSP AND GASIFIER SYSTEM COST

Figure 7-1 shows how the AR gasifier-based IGCC COE change as the AR DSP and gasifier TPC varies from -30% to +30%. Also shown in Figure 7-1 is the reference Case 2a IGCC COE at 145.3 mills/kWh.

Figure 7-1
Sensitivity Analysis – COE vs AR DSP/Gasifier TPC



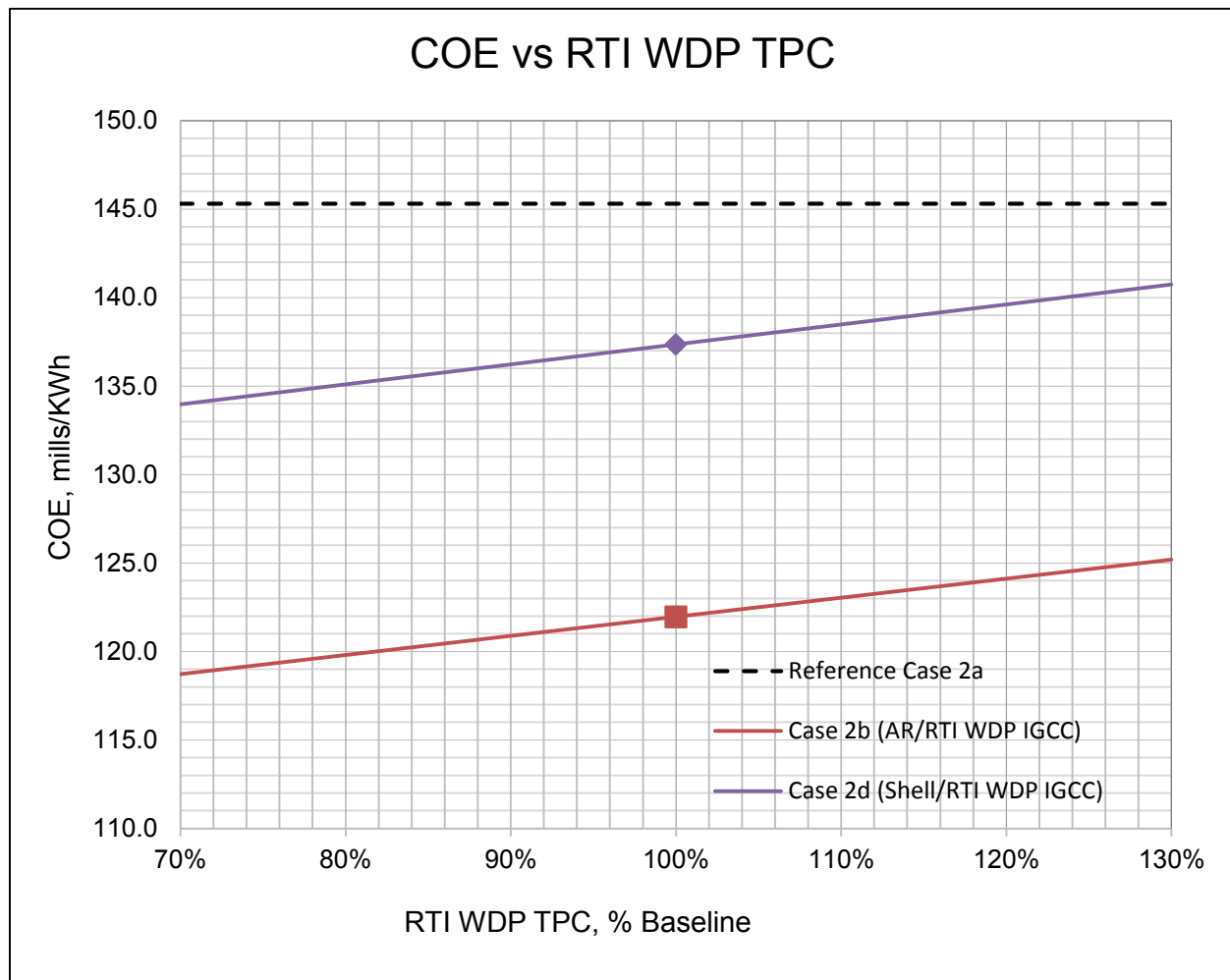
From Figure 7-1, it can be seen that at the high end of the AR advanced gasification TPC (+30%), the COEs for Cases 2b and 2c are still less than the COE for the reference Case 2a with the Shell/two-stage Selexol™ IGCC configuration.

For both the AR gasifier-based IGCC cases, every 5% increase in DSP and gasifier TPC, roughly equivalent to \$16MM in TPC, increases the IGCC COE by 1 mill/kWh.

7.2 RTI WDP SYSTEM COST

Figure 7-2 shows how the RTI WDP-based IGCC (Cases 2b and 2d) COEs change as the RTI WDP system TPC varies from -30% to +30%. The RTI WDP system TPC includes the costs for the RTI WDP, DSRP, WGS and AACRP processes. For reference purposes, the Case 2a IGCC's COE of 145.3 mills/kWh is shown in Figure 7-2 as well.

Figure 7-2
Sensitivity Analysis – COE vs RTI WDP System Cost



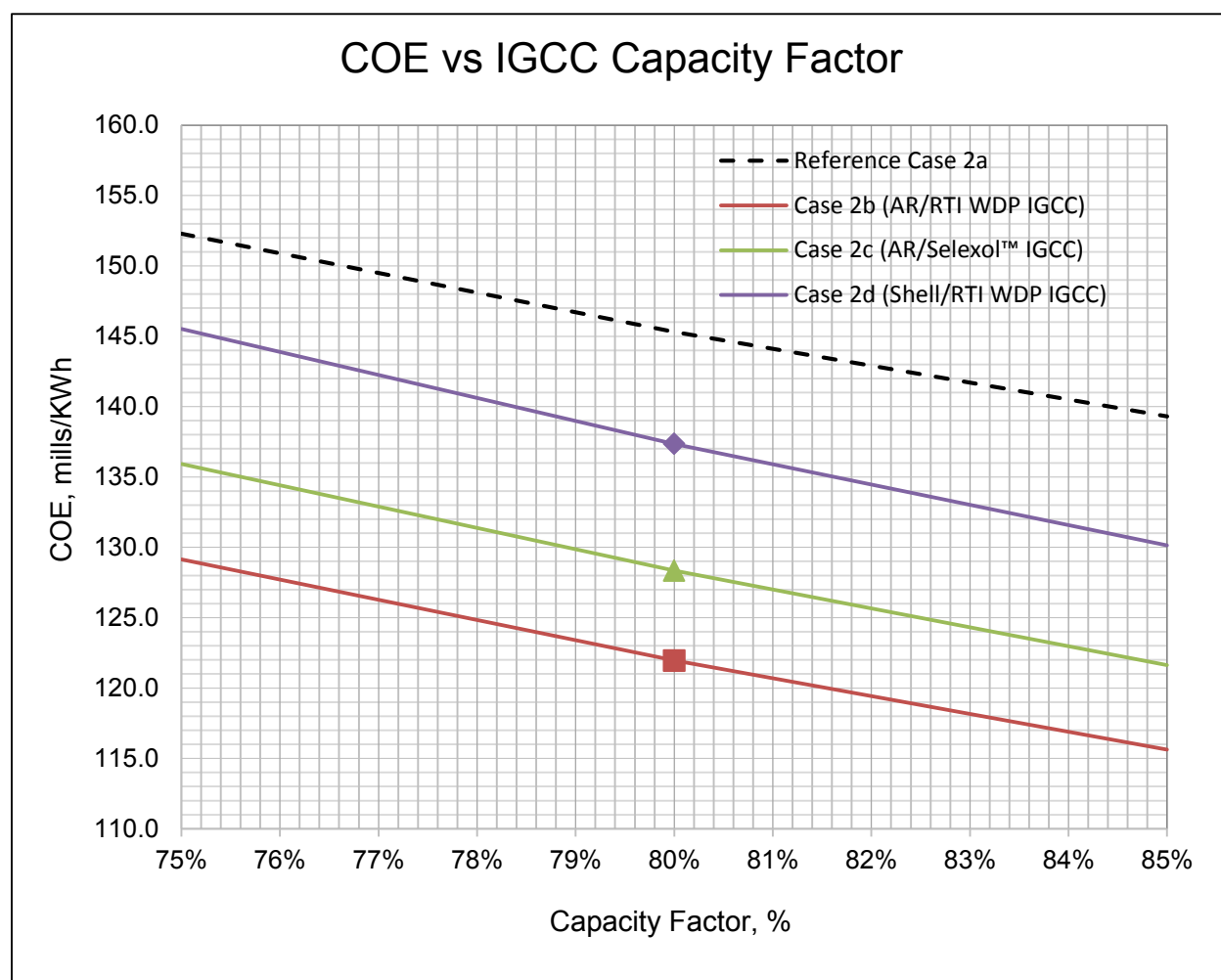
From Figure 7-2, it can be seen that at the high end of the RTI WDP TPC (+30%), the COEs for Cases 2b and 2d are still less than the COE for the reference Case 2a with the Shell/two-stage Selexol™ IGCC configuration.

For both cases, roughly every 9% increase in RTI WDP TPC, equivalent to about \$16MM, increases the IGCC COE by 1 mill/kWh.

7.3 CAPACITY FACTOR

The baseline IGCC plant capacity factor used in this study is 80%. Figure 7-3 shows how the IGCC COE varies with plant CF as it varies from 75% to 85%.

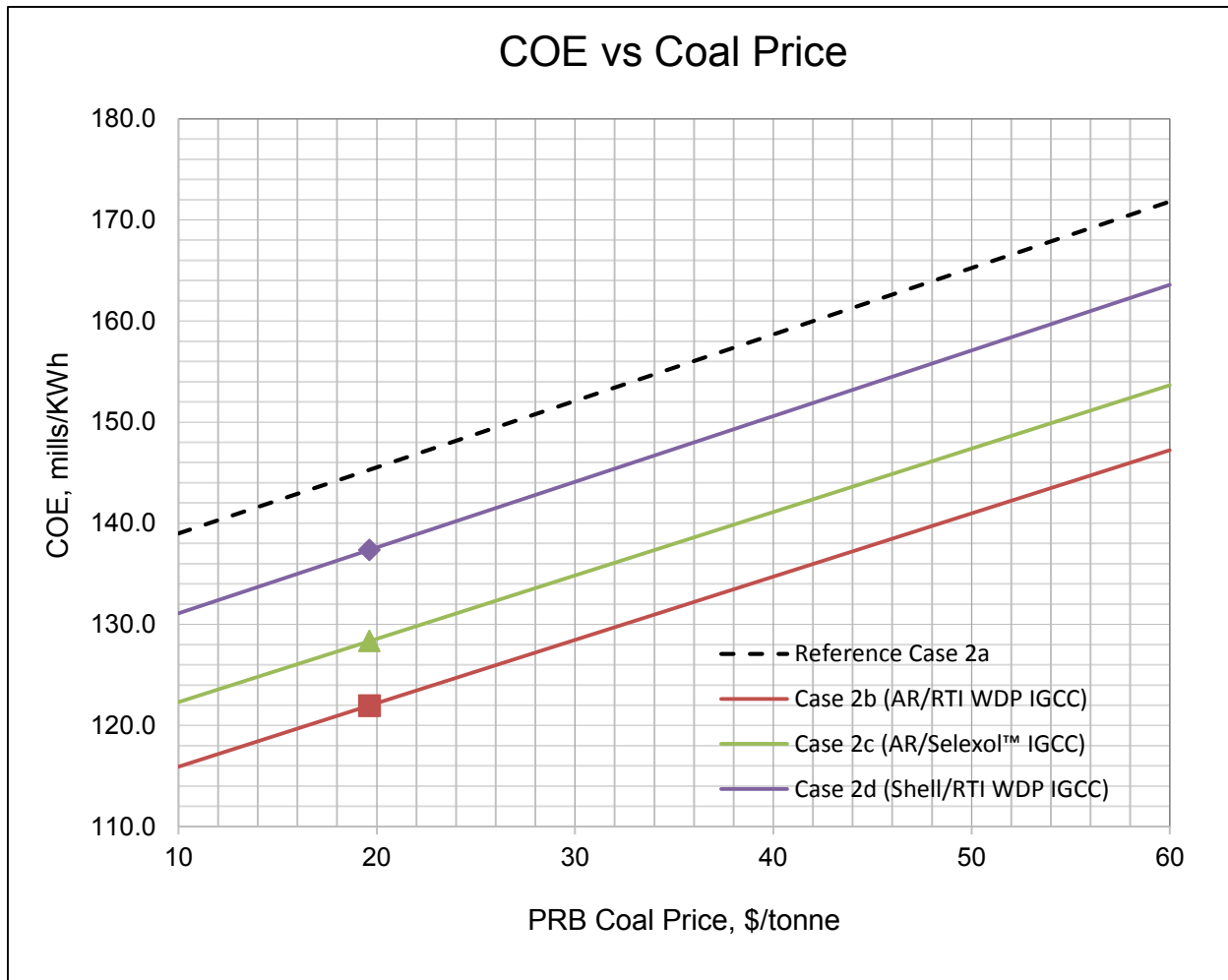
Figure 7-3
Sensitivity Analysis – COE vs IGCC Plant Capacity Factor



7.4 FEEDSTOCK PRICE

The baseline IGCC plant PRB coal feedstock price used in this study is \$19.63/ton. Figure 7-4 shows how the IGCC COE varies with coal price as it varies from \$10/ton to \$60/ton.

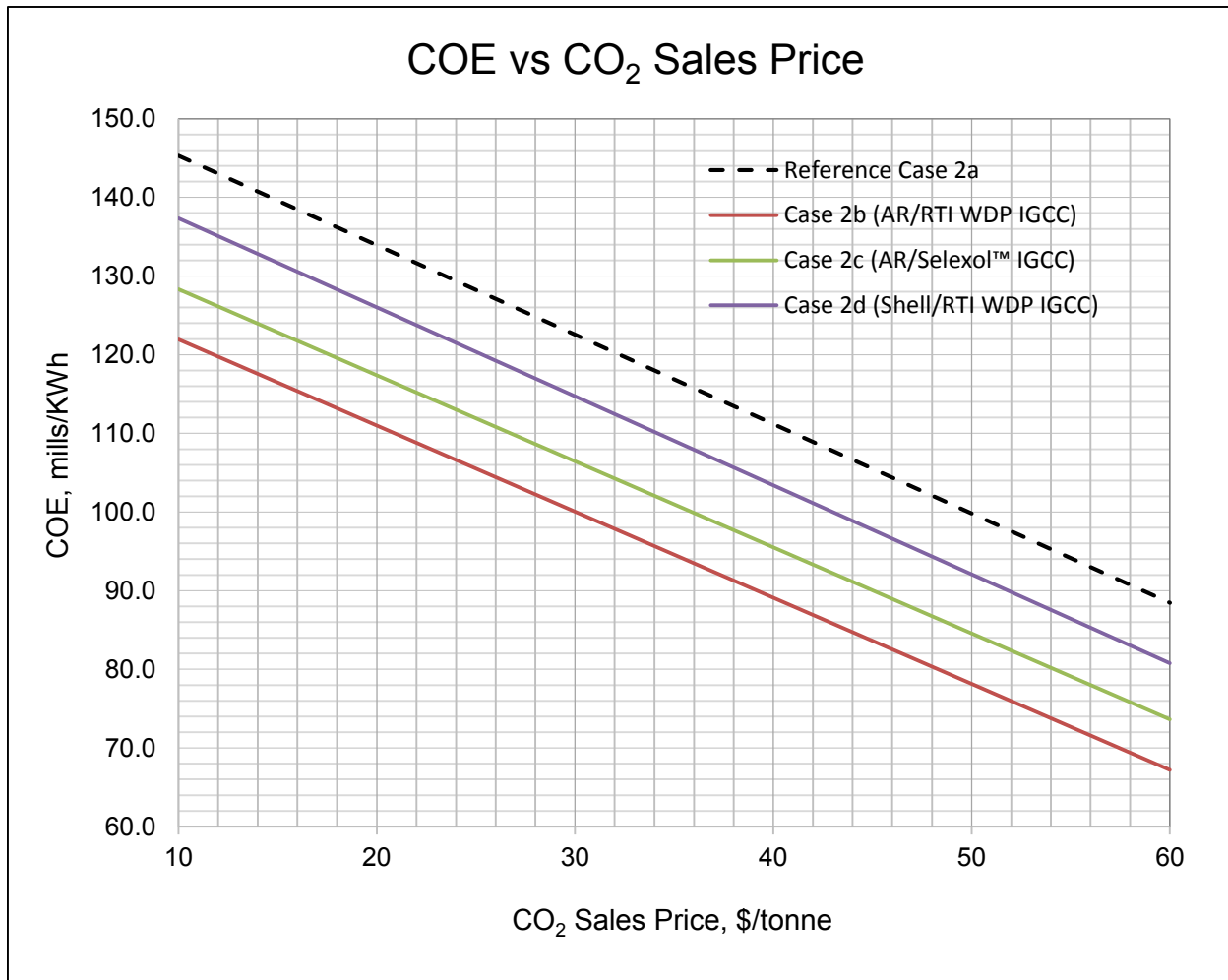
Figure 7-4
Sensitivity Analysis – COE vs Feedstock Price



7.5 CO₂ SALES PRICE

Sensitivity to CO₂ sales at plant gate prices is shown in Figure 7-5. The baseline case assumes that the CO₂ product carries no value (\$0/tonne). The sales price is subsequently varied to a maximum of \$60/tonne to determine its effect on the IGCC plant's COE.

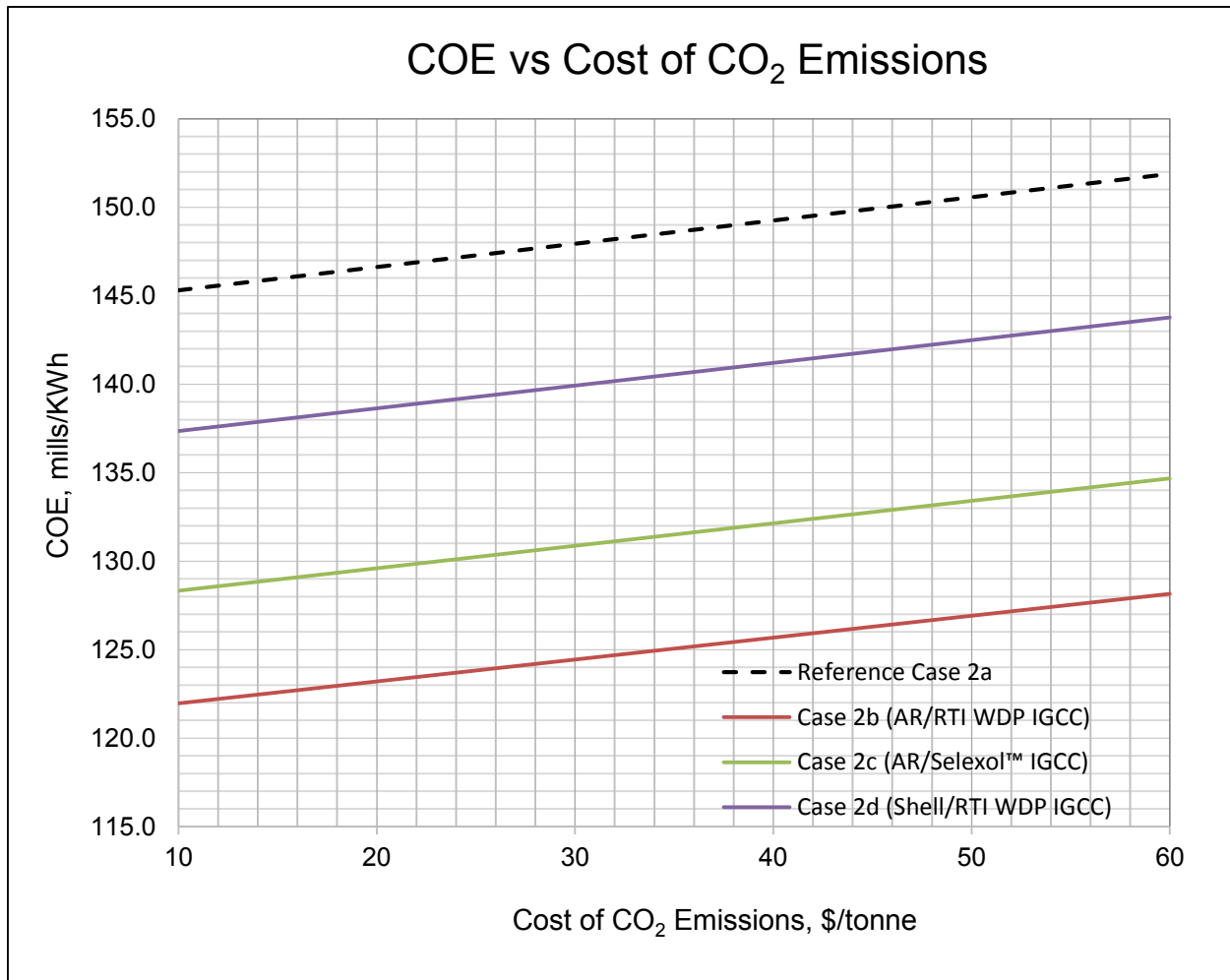
Figure 7-5
Sensitivity Analysis – COE vs CO₂ Sales Price



7.6 COST OF CO₂ EMISSIONS

The sensitivity to CO₂ emissions costs is shown in Figure 7-6. The baseline case assumes that there are no costs associated with venting CO₂ to the atmosphere (\$0/tonne). The cost of CO₂ emissions is subsequently varied to a maximum of \$60/tonne to determine its effect on the IGCC plant's COE.

Figure 7-6
Sensitivity Analysis – COE vs Cost of CO₂ Emissions



The COE are less sensitive to emissions cost than to CO₂ sales price because much more CO₂ is captured by the IGCC plant than is vented (90% vs 10%), hence the magnitude of the COE sensitivity is about 9 times greater to CO₂ sales price than to cost of CO₂ emissions.

8.1 STUDY OBJECTIVE

The objective of this study was to assess the benefits of integrating the advanced technologies offered by RTI and AR for reducing the capital and production costs for hydrogen-rich syngas with near-zero emissions in an IGCC power plant with carbon capture.

The reference case for this study was Case S1B from DOE/NETL's Baseline Studies for power plants in NETL Report 1399. The reasons for selecting this particular case were:

- The Shell gasifier-based IGCC utilizes the pressurized lock hopper dry feed system. This provides a more direct comparison against AR's DSP feed system, which is part of its overall gasification package. Because the reference and AR gasification systems are based on dry coal feed, a more direct comparison can be drawn between the two, with minimal secondary integration impacts associated with differences in ancillary equipment/subsystems.
- Of all the gasification technologies evaluated in the DOE/NETL's Baseline Studies, the Shell-gasification based IGCC has the highest efficiency. If the IGCC utilizing AR/RTI technologies can perform better than the Shell IGCC design in terms of efficiency and cost, it will clearly establish the integrated AR gasification and RTI syngas cleanup system as the best available IGCC technology.
- Shell's gasification technology is considered mature, having already been proven on a commercial scale.

To identify and determine any synergistic advantages of integrating both of these technologies, a total of four case studies were completed. The specific technologies included in each of the four IGCC configurations are identified in the IGCC case study matrix shown in Table 8-1.

Table 8-1
Case Study Matrix for IGCC with CO₂ Capture

	Case 2a ¹	Case 2b	Case 2c	Case 2d
Gasification Technology				
Shell Gasification with Lockhopper-Based Feed System	✓			✓
AR Compact Gasifier with Dry Solids Pump (DSP) Feed System		✓	✓	
Gas Cleanup				
Two-Stage Selexol™ for CO ₂ and Sulfur Removal	✓		✓	
RTI WDP with AACRP		✓		✓
Water Gas Shift				
Sour Shift	✓		✓	
RTI AFWGS		✓		✓
GE 7FB Advanced Gas Turbine	✓	✓	✓	✓
CO ₂ Drying and Compression (to 2,215 psia)	✓	✓	✓	✓

¹ Reference case based on Nexant's benchmark simulation of the DOE/NETL 2010/1399: Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity: IGCC Cases S1B case

All plant configurations include the following process operations: gasifier, quench, water gas shift, low temperature cooling, sulfur removal and CO₂ removal, gas turbine and steam turbine at a greenfield site (Montana, 3,400 ft elevation). Each case uses two GTs for a combined potential gross output of 430 MWe at site elevation. The high site elevation for the reference case impacts the power performance of the IGCC plant compared with more typical TEA results conducted at sea level. Although a full evaluation at sea level was not possible, a preliminary analysis was completed to provide rough approximation of performance at sea level shown in Table 8-2. For this analysis simplifying assumptions about GT output, auxiliary load and costs were made.

Table 8-2
Performance Comparison for Reference Plant at Two Different Elevations

Performance Metric	Reference Case	
	Elevation: 3,400 ft	Elevation: Sea level
Power Production, kWe		
Gas Turbine	429,973	464,000
Steam turbine	224,080	224,080
Auxiliary Power Consumption ¹ , kWe	193,448	193,448
Net Power Output, kWe	460,605	494,632
Total Overnight Cost (2011\$MM)	2,487	2,487
COE, excl CO ₂ TS&M ^{2,3} , mills/kwh	145.3	135.4
COE, incl CO ₂ TS&M ^{2,3} , mills/kwh	166.2	154.8
Cost of CO ₂ Avoided excl CO ₂ TS&M ^{2,3} , \$/ton CO ₂	79.8	67.8
Cost of CO ₂ Avoided incl CO ₂ TS&M ^{2,3} , \$/ton CO ₂	105.1	91.3

¹ Auxiliary loads at sea level have been assumed to be equal to results at 3,400 ft of elevation. The loads will change, but fully evaluating these changes was beyond the scope of this project.

² The costs from the reference case were escalated to 2011\$ based on DOE escalation for GE IGCC case with carbon capture in DOE Report, "Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases".

³ Reference case COE and CO₂ avoided costs at sea level were estimated by RTI.

8.2 RESULTS SUMMARY

Table 8-3 shows a summary comparison of the CAPEX, OPEX, power production, COE and cost of CO₂ avoided for the four cases.

Table 8-3
IGCC Results Summary

Case	Case 2a	Case 2b	Case 2c	Case 2d
IGCC Configuration				
Gasifier	Shell	AR	AR	Shell
Sulfur and CO ₂ Removal	Selexol™	RTI WDP	Selexol™	RTI WDP
CAPEX, \$MM				
Total Installed Cost (TIC)	\$1,521	\$1,242	\$1,309	\$1,444
Total Plant Cost (TPC)	\$2,030	\$1,674	\$1,772	\$1,919
Total Overnight Cost (TOC)	\$2,487	\$2,048	\$2,174	\$2,344
OPEX, \$MM/yr (100% Capacity Factor Basis)				
Fixed Operating Cost (OC _{fix})	\$74.2	\$62.8	\$65.9	\$70.6
Variable Operating Cost, less Fuel (OC _{var})	\$56.9	\$49.1	\$50.5	\$55.2
Fuel (OC _{fuel})	\$50.4	\$48.4	\$48.4	\$50.4
Total OPEX	\$181.5	\$160.2	\$164.8	\$176.2
Power Production, MWe				
Gas Turbine	430.0	429.9	430.2	429.6
Steam Turbine	224.1	211.3	209.3	226.4
Auxiliary Power Consumption	193.4	178.7	177.8	192.2
Net Power Output	460.6	462.5	461.8	463.8
Emissions				
SO ₂ Emissions (lb/MWh _{gross})	0.0063	0.0013	0.0053	0.0014
SO ₂ Emissions (lb/MMBtu)	0.0008	0.0002	0.0007	0.0002
Fuel Rate and Efficiency				
Coal Feed Rate, tpd AR Coal	7,032	6,752	6,752	7,032
Net Efficiency, %	31.32%	32.75%	32.70%	31.53%
Power Generated, MWh/yr (MWh)	4,034,901	4,051,626	4,045,045	4,062,618
TOC CAPEX, \$/kW	5,400	4,428	4,709	5,054
Total OPEX, \$/MWh	44.97	39.55	40.74	43.36
COE, excl CO₂ TS&M, mills/kWh	145.3	122.0	128.3	137.3
COE, incl CO₂ TS&M, mills/kWh	166.2	142.0	148.4	158.1
Cost of CO₂ Avoided excl CO₂ TS&M, \$/ton CO₂	79.8	51.1	59.0	70.0
Cost of CO₂ Avoided incl CO₂ TS&M, \$/ton CO₂	105.1	75.2	83.1	95.0

Table 8-3 clearly quantifies the benefits associated with the integration of AR advanced gasification and RTI's advanced syngas cleanup technologies in an IGCC with carbon capture. Comparison Case 2b, in which these two advanced technologies were integrated achieved the best overall performance of all comparison cases completed in all the following categories:

- Overall plant efficiency: 32.75%
- Capital Cost: \$2,048 MM
- Total operating cost: \$160.2 MM
- COE: 122.0 mills/kWh
- Cost of CO₂ avoided: 51.1 \$/ton CO₂
- SO₂ emissions: 0.0013 lb/MWh_{gross}

The success achieved through integration of these advanced technologies comes from the impressive benefits associated with each individual technology. The AR advanced gasification system, which is comprised of DSP and AR's compact gasifier, had a TPC of \$313 MM (Subaccounts 2.3 plus 4.1). Compared with the TPC for the Shell gasification system of \$510

MM, AR's advanced gasification system only costs 61% as much as Shell's gasification system. This difference represents about a 12 mills/kWh COE advantage for the AR gasification system compared with Shell's gasification system.

RTI's advanced syngas cleanup system also provides a significant reduction in capital costs when compared with a dual-stage Selexol™ system. From Case 2d, the TPC for RTI's advanced syngas cleanup system, which includes syngas desulfurization, sulfur recovery, water-gas shift, CO₂ capture and compression, was \$252 MM (Subaccounts 5A.1, 5A.2, 5A.4, 5A.9, 5B.1 and 5B.2). The TPC for the corresponding system based on Selexol™ was \$341 MM. This results in an \$89 MM capital cost advantage (26% capital cost reduction) and about a 6 mills/kWh advantage in COE.

Additional synergistic benefits are obtained in Case 2b when AR's and RTI's advanced technologies are integrated. In Case 2b, the TPC for the syngas desulfurization, sulfur recovery, water gas shift, CO₂ capture and compression (Subaccounts 5A.1, 5A.2, 5A.4, 5A.9, 5B.1 and 5B.2) is \$246 MM. The corresponding system in Case 2a (Reference case) has a TPC of \$341 MM which results in a \$95 MM capital cost advantage for WDP over the Selexol™ process. This represents an additional capital cost advantage for integration of both advanced technologies of \$6 MM.

From these conclusions, the integration of AR advanced gasification and RTI advanced syngas cleanup processes results in the following benefits when compared with the best available commercial technologies:

- ~ 1.4% net efficiency improvement,
- ~18% reduction in capital cost per kW generated,
- ~12% reduction in the operating cost per kW generated, and
- ~ 75-79% reduction in sulfur emissions.

Because of the significant economic and efficiency benefits, market forces will drive commercial implementation of these technologies to realize the economic benefits. The fact that integration of these two technologies optimizes the benefits will also drive integrated deployment of these technologies. The fact that the RTI advanced syngas cleanup process also reduces sulfur emissions creates a condition where market forces will also drive the implementation of cleaner technologies.

The sensitivity analysis completed shows that the benefits in COE for AR's gasification and RTI's WDP technologies still exists for a 30% increase in the total plant cost for these technologies. All of these factors indicate that when ready for commercial deployment, these technologies will be favored over existing commercial technologies.

°F	Degree Fahrenheit
AGR	Acid Gas Removal
AACRP	Advanced Amine CO ₂ Removal Process
AOI	Area of Interest
AR	Aerojet Rocketdyne
Ar	Argon
ASU	Air Separation Unit
B/L	Battery Limit
BEC	Bare Erected Cost
BFD	Block Flow Diagram
BFW	Boiler Feed Water
BOP	Balance of Plant
Btu	British Thermal Unit
CAPEX	Capital Expenditure
CCF	Capital Charge Factor
CF	Capacity Factor
CH ₄	Methane
Circ	Circulating
CO	Carbon Monoxide
CO ₂	Carbon Dioxide
COE	Cost of Electricity
COS	Carbonyl Sulfide
CTL	Coal-to-Liquids
CTG	Combustion Turbine Power Generation
CW	Cooling Water
DBT	Dry Bulb Temperature
DCS	Distributed Control System
DOE	U.S. Department of Energy
DSP	Dry Solids Pump
DSRP	Direct Sulfur Recovery Process
EPC	Engineering, Procurement and Construction
EPRI	Electric Power Research Institute
EPRI	Electric Power Research Institute
FCC	Fluid Catalytic Cracking
FO	Fuel Oil
FOA	Funding Opportunity Announcement
ft	feet

GE	General Electric
H ₂	Hydrogen
H ₂ O	Water
H ₂ S	Hydrogen Sulfide
Hg	Mercury
HGCU	Hot Gas Clean Up
HHV	Higher Heating Value
HMB	Heat and Material Balance
HP	High Pressure
hr	Hour
HRSG	Heat Recovery Steam Generator
I & C	Instrumentation & Control
IGCC	Integrated Gasification Combined Cycle
IOU	Investor Owned Utility
IP	Intermediate Pressure
kWe	Kilowatt electric
kWh	kilowatt hour
lb	Pound Mass
LHV	Lower Heating Value
LP	Low Pressure
max	Maximum
ME	Major Equipment
MEC	Major Equipment Cost
min	Minimum
Misc	Miscellaneous
MM	million
MU	Makeup
MWe	Megawatt electric
MWh	megawatt hour
N ₂	Nitrogen
NETL	National Energy Technology Laboratory
NO _x	Oxides of Nitrogen
NSPS	New Source Performance Standards
O&M	Operating and Maintenance
O ₂	Oxygen
OPEX	Operating Expenditure
OSBL	Outside Battery Limit
PC	Pulverized Coal
PFB	Pressurized Feed Bin
PFD	Process Flow Diagram

PM	Particulate Matter
ppmv	Parts per Million by Volume
ppmW, ppmw	Parts per Million by Weight
PRB	Powder River Basin
PSFM	Power Systems Financial Model
psi	Pounds Per Square Inch
psia	Pounds Per Square Inch, absolute
psig	Pounds Per Square Inch, gauge
QGESS	Quality Guidelines for Energy System Studies
RTI	Research Triangle Institute
SC	Supercritical
scfm	Standard Cubic Feet per Minute
SO ₂	Sulfur Dioxide
SOPO	Statement of Project Objectives
SRU	Sulfur Recovery Unit
STG	Steam Turbine Power Generation
T&S	Transportation and Storage
TDC	Total Direct Cost
TEA	Techno-Economic Analysis
TFC	Total Field Cost
TG	Turbine Generator
TGTU	Tail Gas Treatment Unit
TIC	Total Installed Cost
TOC	Total Overnight Cost
TPC	Total Plant Cost
tpd	tons per day
US, USA	United States of America
vol%	Percentage by Volume
WBT	Wet Bulb Temperature
WDP	Warm Gas Desulfurization Process
WGS	Water Gas Shift
WHR	Waste Heat Recovery
WT	Waste Treatment

11. Appendix B

Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production

Task 3: Techno-Economic Analysis for Crude Methanol Production from Coal and Natural Gas Application

Submitted by

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PREFACE

Although the same directional improvement in cost and technical information that fueled our enthusiasm for the results from a sister study for Integrated Gasification Combined Cycle (IGCC) applications with 90% carbon capture is present in this techno-economic analysis study, our excitement about this study was more focused on the application of our advanced technologies for the production of chemicals. Gasification, and in particular coal gasification, has been and is expected to continue to be more widely used globally for chemicals production than for IGCC power generation (at least for the near- to mid-term).

The primary reason we see a brighter future for gasifier-based chemical production comes from the fact that chemicals production, in general, has higher relative value creation potential than electric power generation and market forces drive chemicals production to be extremely competitive (as compared to public utilities that may have multiple alternative drivers such as overall grid reliability and fuel diversity). The revenue from chemical production comes from the profit margin between the market price of the chemical and its actual production cost. Since there are limited options to influence the market price, chemical producers are continuously seeking to lower production costs. In support of this goal, they look for the lowest cost technologies, which has proven to be a major concern area for many conventional gasification-based plants and projects. Chemical producers are also very proactive in terms of operation of their facilities, emphasizing the importance of interactive controls to keep the plant operating at peak on-line performance and the incredible importance of continuous improvement to increase availability and overall performance of their facilities. This operating philosophy/strategy, coupled with inherently aligned capabilities and resources to support chemical-based technologies such as gasification, provides a more natural home for gasification processes and leads to systematic improvements that drive the technologies to be more reliable and more effective. For example, in China the impact of the decision to use coal gasification to support a large portion of the country's chemical industry has resulted in significant reduction in costs and has driven technical innovation. We believe that emerging technologies in gasification-based applications will benefit significantly from projects that are chemical production based.

In this study, a 90% overall carbon capture requirement attempts to address greenhouse gas emissions concerns. Because of this constraint, the costs of the coal-to-methanol (CTM) plants in this study were required to include a dedicated natural gas power plant with 90% carbon capture (to supply all auxiliary power needs for the CTM plant) in addition to carbon capture for the CTM plant itself. This has several very negative impacts. If any excess power is generated, this excess power would likely be sold to the grid at a discount because any negotiated power purchase agreement would exploit the fact that the power production is an inherent by-product of the methanol process. Thus any excess power generation enabled by use of our more efficient advanced technologies has essentially no financial value, whereas a decision in the base case to allow power needs to be purchased across the fence from the grid (even if as carbon-neutral power) would have enabled the net power generated by our advanced technologies to be used to offset power purchases and reduce the methanol required selling price (RSP). The requirement to include power generation processes with 90% carbon capture also substantially increases the overall capital costs of the modeled CTM plants. As the capital cost represents over 60% of the methanol RSP, the cost of the power generation system and supporting subsystems adversely inflate the overall capital cost of the CTM plant and methanol RSP and somewhat diminish the relative net impact of utilizing our advanced technologies. Unfortunately sectioning out the capital cost of the power generating equipment and the associated increase in capital cost of the supporting systems is not possible in the current study, because of integration between the process heat recovery and power generation system. If a company interested in installing a CTM

plant attempts to use the cost results from this study, it should be noted that the capital costs reported will be significantly higher than that of a typical commercial plant without the inclusion of an associated power plant with carbon capture.

A more viable approach would have been to fix a cost for electricity, which included carbon capture costs. In this manner, the CTM plant could address its carbon foot print and the electric power provider would be responsible for its own carbon foot print. Coupled with sensitivity analysis, this could help identify cost targets that would start making carbon capture economically viable for industry as a whole. It would also help identify the most important areas for research and development with quantifiable targets.

In spite of this power neutral constraint, the results from this study show that Aerojet Rocketdyne's and RTI's emerging technologies provide both cost and sulfur and carbon dioxide emissions reductions for the chemical production of methanol from coal. Based on the current technology roadmaps, both of these technologies will be available for commercial deployment within the next few years. The availability and benefits of these technologies will result in market forces that favor the implementation of these technologies, helping to reduce costs and emissions from coal utilization for both chemical and power production.

December 2016

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ABSTRACT

As a follow-on from a previous Department of Energy/National Energy Technology Laboratory (DOE/NETL) funded techno-economic study for using hydrogen-rich for Integrated Gasification Combined Cycle (IGCC) with > 90% carbon capture, this techno-economic report represents one of two studies evaluating integration of advanced technologies in Coal-to-Liquids (CTL) applications. This techno-economic study focuses on Coal-to-Methanol (CTM) with >90% carbon capture. The two advanced technologies evaluated included Aerojet Rocketdyne's (AR's) advanced gasification system (currently being offered by Gas Technology Institute as R-GAS™) and RTI's advanced warm syngas cleanup technology. The reference case for this study was Case 2 from "*Base line Analysis of Crude Methanol Production from Coal and Natural Gas*, Revised July 29, 2014, DOE/NETL, 341/020514." Case 2 from this study was the only case that included 90% carbon capture. A Natural Gas Combined Cycle (NGCC) with the necessary systems was also included to ensure 90% carbon capture while making the CTM plant power neutral. In addition to the reference case, three additional cases were developed. For the first study, both advanced technologies were integrated into the CTM plant. In the remaining two studies, AR's and RTI's advanced technologies were integrated into the CTM plant individually to help evaluate synergistic benefits.

From these techno-economic case studies, each advanced technology did individually result in reduction of the required selling price (RSP) of the methanol, but it was the integrated case with both advanced technologies that offered the most benefits. In comparison with the reference case, the benefits of the integrated case with both advanced technologies included:

- 17-19% reduction in the RSP of the methanol product for both the commercial fuel and loan guarantee project financial structures,
- 22% reduction in CAPEX,
- 13% reduction in OPEX,
- ~24 MWe of surplus electrical power,
- 99% reduction in sulfur emissions, and
- 13-15% reduction in CO₂ emissions.

From sensitivity analysis, there is still a significant reduction in the methanol RSP for the CTM plant with the AR gasification and the RTI advanced syngas cleanup technologies even when the cost of these technologies had been increased by 30%. This unique situation, in which the installation of advanced technologies improves process economics and lowers sulfur and CO₂ emissions, will favor the deployment of these technologies over existing commercial technologies when these technologies are ready for commercial deployment within the next few years.

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EXECUTIVE SUMMARY

With U.S. Department of Energy National Energy Technology Laboratory (DOE/NETL) funding, Research Triangle Institute (RTI), Aerojet Rocketdyne (AR) and Nexant completed a set of techno-economic analyses to assess the potential benefits associated with integrating the emerging technologies being offered by RTI and AR for reducing the production costs for hydrogen-rich syngas with near-zero emissions from coal gasification for both integrated gasification combined cycle (IGCC) and Coal-to-Liquids (CTL) applications.

RTI's Warm Gas Desulfurization Process (WDP) and AR's Advanced Gasification System are emerging technologies that will enable commercial deployment of gasification through reduced costs, increased efficiency and greater contaminant removal within the next five years. AR's compact gasifier is an entrained-flow, dry-feed gasification system that utilizes AR's proprietary dry solids feed pump (DSP). Due to their novel designs, AR's compact gasifier and dry solids pump significantly reduce the size and amount of equipment, which is projected to reduce the cost of the gasification system by approximately half.

RTI's advanced syngas cleanup process technologies consist of the following components: RTI's WDP, RTI's direct sulfur recovery process (DSRP), and RTI's advanced fixed-bed water gas shift process (AFWGS). An activated amine carbon dioxide recovery process (AACRP) has also been integrated with RTI's advanced warm syngas cleanup process, because through this integration contaminant removal equivalent to that of Rectisol[®] can be achieved at a fraction of the cost. With RTI's 50 MWe pre-commercial test plant having successfully demonstrated the technology at Tampa Electric Company's Polk Power Station, RTI's WDP advanced syngas cleanup technology will be ready for commercial deployment in the next year.

To fully assess the benefits of each advanced technology, a total of four techno-economic analyses, which are shown in Table E-1, were completed which included >90% carbon capture. The first of these techno-economic analyses was a reference case, which was Case 2 from DOE/NETL's "*Baseline Analysis of Crude Methanol Production from Coal and Natural Gas*, Revised July 29, 2014, DOE/NETL. 341/020514" (Crude Methanol Study). This case was selected because it was the only case study which offered >90% carbon capture. In addition, this case included a natural gas combined cycle (NGCC) plant with > 90% carbon capture to make the CTM plant power neutral and achieve the >90% total carbon capture.

Based on the four cases shown in Table E-1, Nexant developed heat and mass balances for the four cases. Then applying the assumptions and methodology established in DOE/NETL's Series of Quality Guidelines for Energy Systems Studies (QGESS), Nexant developed factored cost estimates for each case. Nexant used the reference case to validate and fine tune their techno-economic methodology. Using DOE/NETL's methodology, AR and RTI developed the necessary heat and mass balances, equipment costs, and utility and consumable needs and costs for their advanced technology processes based on their pilot plant and demonstration experience. The results from these case studies are summarized in Table E-2. The calculation of the required selling price (RSP) for the methanol product was calculated based on financing typical for commercial fuel projects and loan guarantee projects. When these cost estimates had been

completed, Nexant also conducted sensitivity analyses for capital cost, capacity factor, natural gas price, coal price, electricity price, capacity charge factor, CO₂ sales prices and cost of CO₂ emissions.

Table E-1
Case Study Matrix for Crude Methanol Production Plant with CO₂ Capture

	Case 3a ¹	Case 3b	Case 3c	Case 3d
Gasification Technology				
Shell Gasification with Lockhopper-Based Feed System	✓			✓
AR Compact Gasifier with Dry Solids Pump (DSP) Feed System		✓	✓	
Gas Cleanup				
Rectisol® for CO ₂ and Sulfur Removal ²	✓		✓	
RTI WDP with AACRP		✓		✓
Water Gas Shift				
Sour Shift	✓		✓	
AFWGS		✓		✓
Methanol Production	✓	✓	✓	✓
NGCC Power Generation with Fluor Econoamine CO ₂ Capture	✓	✓	✓	✓
CO ₂ Drying and Compression (to 2,215 psia)	✓	✓	✓	✓

¹ Reference case based on Nexant's benchmark simulation of the Crude Methanol Study Case 2

² Rectisol® removes H₂S and CO₂. Additional trace contaminant cleanup technologies will be included as defined by DOE/NETL baseline studies

Table E-2 clearly quantifies the benefits associated with the integration of AR's advanced gasification and RTI's advanced syngas cleanup technologies in the CTM plant with carbon capture. Comparison Case 3b, in which these two advanced technologies were integrated achieved the best overall performance of all comparison cases with the following benefits compared to the reference case:

- 17-19% reduction in the required selling price (RSP) of the methanol product for both the commercial fuel and loan guarantee project financial structures,
- 22% reduction in CAPEX,
- 13% reduction in OPEX,
- ~24 MWe of surplus electrical power,
- 99% reduction in sulfur emissions, and
- 13-15% reduction in CO₂ emissions.

Because of the significant economic and efficiency benefits, market forces will drive commercial implementation of these technologies to realize these economic benefits. The fact that integration of these two technologies optimizes the benefits will also drive integrated deployment of these technologies. The fact that the RTI advanced syngas cleanup process also reduces sulfur emissions creates a condition where market forces will also drive the implementation of cleaner technologies.

Table E-2
CTM Results Summary

Case	Case 3a	Case 3b	Case 3c	Case 3d
CTM Configuration				
Gasifier	Shell	AR	AR	Shell
Sulfur and CO ₂ Removal	Rectisol®	RTI WDP	Rectisol®	RTI WDP
CAPEX, \$MM				
Total Installed Cost (TIC)	\$3,581	\$2,732	\$2,897	\$3,386
Total Plant Cost (TPC)	\$4,802	\$3,707	\$3,950	\$4,513
Total Overnight Cost (TOC)	\$5,892	\$4,567	\$4,857	\$5,548
OPEX, \$MM/yr (100% CF Basis)				
Fixed Operating Cost (OC _{fix})	\$162.9	\$128.0	\$135.6	\$153.7
Variable Operating Cost, less Fuel (OC _{var})	\$103.9	\$95.3	\$87.9	\$111.2
Coal Feedstock (OC _{Coal})	\$259.2	\$244.8	\$244.8	\$259.2
Natural Gas Feedstock (OC _{NG})	\$85.4	\$77.6	\$78.4	\$85.4
Power Import/(Export)	\$0.4	(\$14.1)	(\$4.0)	(\$12.9)
Total OPEX	\$611.8	\$531.4	\$542.7	\$596.6
Methanol Production, mtpd (100% CF Basis)				
Gross Methanol Production	10,218	10,079	10,204	10,100
Rectisol® Methanol Makeup	8	0	6	0
Net Methanol Production	10,210	10,079	10,198	10,100
NGCC Plant Power Generation, MWe				
Gas Turbine	127.1	127.9	127.8	127.1
Steam Turbine	264.7	239.2	199.1	292.8
Total Power Generation	391.8	367.1	326.8	419.9
Auxiliary Consumption, MWe				
<i>Process Plant</i>				
Coal Handling and Drying	17.4	16.4	16.4	17.7
Gasification and ASU	185.3	163.6	165.1	184.3
Process AGR	42.5	42.6	31.7	39.1
Process CO ₂ Compression	65.2	64.8	52.4	75.6
Methanol Synthesis	27.3	4.7	5.0	23.2
Process Plant BOP	22.5	20.5	20.2	23.0
<i>Total Process Plant Auxiliary Consumption</i>	<i>360.2</i>	<i>312.5</i>	<i>290.9</i>	<i>362.9</i>
<i>NGCC Plant</i>				
NGCC Auxiliary Consumption	22.4	23.5	17.7	28.4
NGCC PCC	3.9	4.1	4.1	3.9
NGCC PCC CO ₂ Compression	6.0	w/Process	6.4	w/Process
<i>Total NGCC Auxiliary Consumption</i>	<i>32.3</i>	<i>27.6</i>	<i>28.2</i>	<i>32.2</i>
Total Auxiliary Consumption	392.6	340.1	319.1	395.1
Net Power Generation/(Consumption)	(0.8)	26.9	7.7	24.8
Fuel Rate and Efficiency				
Coal Feed Rate, tpd AR Coal	19,418	18,339	18,339	19,418
Natural Gas Feed Rate, MMBtu/day	38,152	34,675	35,051	38,152
Total Thermal Input In (LHV), MMBtu/day	354,934	333,984	334,323	354,934
Methanol Fuel Thermal Output (LHV), MMBtu/day	188,547	185,730	188,421	185,877
Power Output, MMBtu/day	(63)	2,207	634	2,029
Thermal Efficiency (LHV), %	53.1%	56.3%	56.5%	52.9%
CO₂ Capture				
CO ₂ Capture Product, tpd	19,413	18,201	17,576	20,021
CO ₂ Vented, tpd	2,966	2,359	2,833	2,538
Carbon Emitted as CO ₂ , % Carbon in Fuel	7.9%	6.6%	7.9%	6.7%

Table E-2 (cont'd)
CTM Results Summary

Case	Case 3a		Case 3b		Case 3c		Case 3d	
Methanol RSP								
Methanol RSP, excl CO₂ TS&M, \$/ton^{BE}	424.2	500.7	347.3	407.3	359.5	422.7	408.9	481.7
, incl CO₂ TS&M, \$/ton^{BE}	441.5	517.9	363.6	423.7	375.1	438.3	426.8	499.6
RSP Components Details, \$/gal								
Capital ^B	0.90	1.15	0.70	0.91	0.74	0.95	0.85	1.10
Fixed O&M	0.15		0.12		0.12		0.14	
Variable O&M	0.08		0.08		0.07		0.09	
Coal	0.21		0.20		0.20		0.21	
Natural Gas	0.07		0.06		0.06		0.07	
Power	0.00		(0.01)		(0.00)		(0.01)	
CO ₂ TS&M	0.06		0.05		0.05		0.06	
Total^B	1.46	1.72	1.21	1.40	1.24	1.45	1.41	1.66
Cost of CO₂ Captured, \$/tonne CO₂^{BC}	18.8	20.9	18.9	21.1	18.7	21.0	19.0	21.1
Cost of CO₂ Avoided, \$/tonne CO₂^{BD}	29.8	31.9	29.9	32.1	29.7	32.0	30.0	32.1

The sensitivity analysis completed shows that the benefits in methanol RSP for AR's gasification and RTI's WDP technologies still exists for a 30% increase in the total plant cost for these technologies. All of these factors indicate that when ready for commercial deployment, these technologies will be favored over existing commercial technologies.

Section 1 Introduction

1.1 BACKGROUND

With U.S. Department of Energy National Energy Technology Laboratory (DOE/NETL) funding from Cooperative Agreement DE-FE0012066, Research Triangle Institute (RTI), Aerojet Rocketdyne (AR), and Nexant formed a team to evaluate the cost and performance benefits associated with integration of the advanced gasification technologies being developed by AR and the warm gas desulfurization process being developed by RTI for hydrogen-rich syngas production for Integrated Gasification Combined Cycle (IGCC) and Coal-To-Liquids (CTL) applications with $\geq 90\%$ carbon capture.

RTI's Warm Gas Desulfurization Process and AR's Advanced Gasification System are next generation technologies that will enable commercial deployment of gasification through improved efficiency, greater contaminant removal, and lower costs within the next five years. AR's compact gasifier is an entrained-flow, dry-feed gasification system that utilizes a dry solids pump (DSP) as the coal feed system. Due to its novel design, the gasifier is drastically smaller (90% size reduction) than a conventional gasifier. Used in conjunction with the DSP, which replaces a large and costly lockhopper-based pressurized coal feed system, the overall compact gasification system is projected to cost as little as 50% of that of a conventional gasification system.

RTI's advanced syngas cleanup process consists of the following components: RTI's Warm Gas Desulfurization Process (WDP), RTI's Direct Sulfur Recovery Process (DSRP), and Advanced Fixed-bed Water Gas Shift (AFWGS). An activated amine carbon dioxide (CO₂) recovery process (AACRP) is also integrated with RTI's advanced syngas cleanup process to effectively remove CO₂. This combination effectively decouples the separation of the sulfur and CO₂ enabling more efficient removal of these contaminants at lower cost. Because this combination is also able to produce syngas with contaminant concentrations similar to a Rectisol® process, RTI's advanced syngas cleanup process offers the potential for significant reduction in contaminant emissions at lower cost than commercially available technologies. With RTI's 50 MWe demonstration plant project (DE-FE0000489) currently demonstrating the technology at Tampa Electric Company's Polk Power Station, RTI's advanced syngas cleanup technology will be ready for commercial deployment within the next year.

1.2 STUDY OBJECTIVES

The overall objective of this project is to assess the potential of integrating the technologies offered by RTI and AR in order to reduce the capital and production costs for hydrogen-rich syngas with near-zero emissions from coal gasification for both power and CTL productions with carbon capture. To fully evaluate the integration of these advanced technologies, separate studies were completed for hydrogen-rich syngas production for power and CTL applications. This report documents the study completed for Coal-to-Methanol (CTM). Separate reports have also been prepared for studies. One evaluates the use of these advanced technologies for an alternative CTL application in which Fischer-Tropsch liquids are produced. The second evaluates the integration of the technologies for the power production for an IGCC.

The specific case studies completed provide a comparison of each advanced technology individually and as an integrated plant with a reference plant using commercially available technologies. All comparison studies conducted for this report capture at least 90% of the carbon in the feedstock as either CO₂ for storage or in the liquid product.

2.1 DESIGN REFERENCES

The reference plant design used for the coal-to-liquids study was selected from “Baseline Analysis of Crude Methanol Production from Coal and Natural Gas, Revised July 29, 2014, DOE/NETL. 341/020514” (Crude Methanol Study). The reference coal-to-methanol (CTM) case used for comparison against the AR gasification and RTI advanced syngas cleanup and AACRP systems was the Crude Methanol Study’s Case 2 CTM plant with CO₂ capture and sequestration. The Crude Methanol Study, along with the following DOE/NETL’s Series of Quality Guidelines for Energy Systems Studies (QGESS):

- “Specifications for Selected Feedstocks, January 2012, DOE/NETL. 341/011812”,
- “Process Modeling Design Parameters, January 2012, DOE/NETL. 341/081911”,
- “CO₂ Impurity Design Parameters, January 2012, DOE/NETL. 341/011212”,
- “Detailed Coal Specifications, January 2012, DOE/NETL-401/01211”,
- “Cost Estimation Methodology for NETL Assessments of Power Plant Performance, April 2011, DOE/NETL. 2011/1455”,
- “Capital Cost Scaling Methodology, January 2013, DOE/NETL. 341/013113”, and
- “Fuel Prices for Selected Feedstocks in NETL Studies, November 2012, DOE/NETL 341/11212”

contain a comprehensive set of design bases and assumptions for the methanol production plant, as well as reference costs and economic evaluation guidelines that were used to complete the techno-economic analyses (TEAs) in this report.

Additionally, “Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2, November 2010, DOE/NETL. 2010/1397” (NETL Report 1397) contains a natural gas combined cycle (NGCC) design with CO₂ capture via Fluor’s Econoamine FG+ process. Its performance provides the basis for the design of the methanol synthesis plant’s power cycle and post-combustion CO₂ capture.

DOE/NETL’s report, “Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases, August 2012, DOE/NETL-341/082312” (NETL Report 341/082312), was also used as the reference to develop the updated NGCC capital and operating cost estimates in June 2011 dollars.

2.2 CASE CONFIGURATIONS

To identify and determine any synergistic advantages of integrating both AR’s advanced gasification and RTI’s advanced syngas cleanup process, a total of four case studies were completed. Case 2 from the Crude Methanol Study was selected as the reference case. The most promising case study was anticipated to be the CTM plant with CO₂ capture integrating both

AR's advanced gasification and RTI's advanced syngas cleanup process. The final two CTM cases evaluated utilizing either the AR advanced gasification system or RTI advanced syngas cleanup process on a standalone basis.

The four crude methanol production plant configurations are identified in the case study matrix shown in Table 2-1.

Table 2-1
Case Study Matrix for Crude Methanol Production with CO₂ Capture

	Case 3a ¹	Case 3b	Case 3c	Case 3d
Gasification Technology				
Shell Gasification with Lockhopper-Based Feed System	✓			✓
AR Compact Gasifier with Dry Solids Pump (DSP) Feed System		✓	✓	
Gas Cleanup				
Rectisol® for CO ₂ and Sulfur Removal ²	✓		✓	
RTI WDP with AACRP		✓		✓
Water Gas Shift				
Sour Shift	✓		✓	
AFWGS		✓		✓
Methanol Production	✓	✓	✓	✓
NGCC Power Generation with Fluor Econoamine CO ₂ Capture	✓	✓	✓	✓
CO ₂ Drying and Compression (to 2,215 psia)	✓	✓	✓	✓

¹ Reference case based on Nexant's benchmark simulation of the Crude Methanol Study Case 2

² Rectisol® removes H₂S and CO₂. Additional trace contaminant cleanup technologies will be included as defined by DOE/NETL baseline studies

2.2.1 Case 3a: Reference Shell Crude Methanol Production Plant with Rectisol®-Based Acid Gas Removal (AGR)

Based on DOE/NETL's recommendation, the reference case was selected from the CTM cases in the Crude Methanol Study. The choice of Case 2 from this study as the reference case is straightforward since it is the only CTM case that includes CO₂ capture with sequestration. In Case 2, a Shell-gasification system is used to generate enough syngas to produce approximately 10,000 metric tons of methanol per day (3,326,000 gal/day based on 332.6 gal/tonne). This plant size is considered large scale but typical for the current design of new methanol plants. Table 2-2 provides a more detailed description of Case 2 of the Crude Methanol Study.

The two-stage Rectisol® process is designed for sulfur and CO₂ removal with a target of 90% for CO₂ removal. The amine capture process for flue gas from the NGCC is also designed for 90% CO₂ removal. The 90% design for both CO₂ point sources results in a total carbon capture rate of >90% for the overall plant.

Both the gasification and methanol synthesis processes generate a large amount of heat that is recovered in the form of steam for process requirements and power generation. Additional power is generated via a NGCC system. In combination with the associated steam turbines, the total power generated from the methanol plant's power cycle in the reference case is equal to its total estimated auxiliary loads i.e. the plant's operations is power neutral, with zero export/import power.

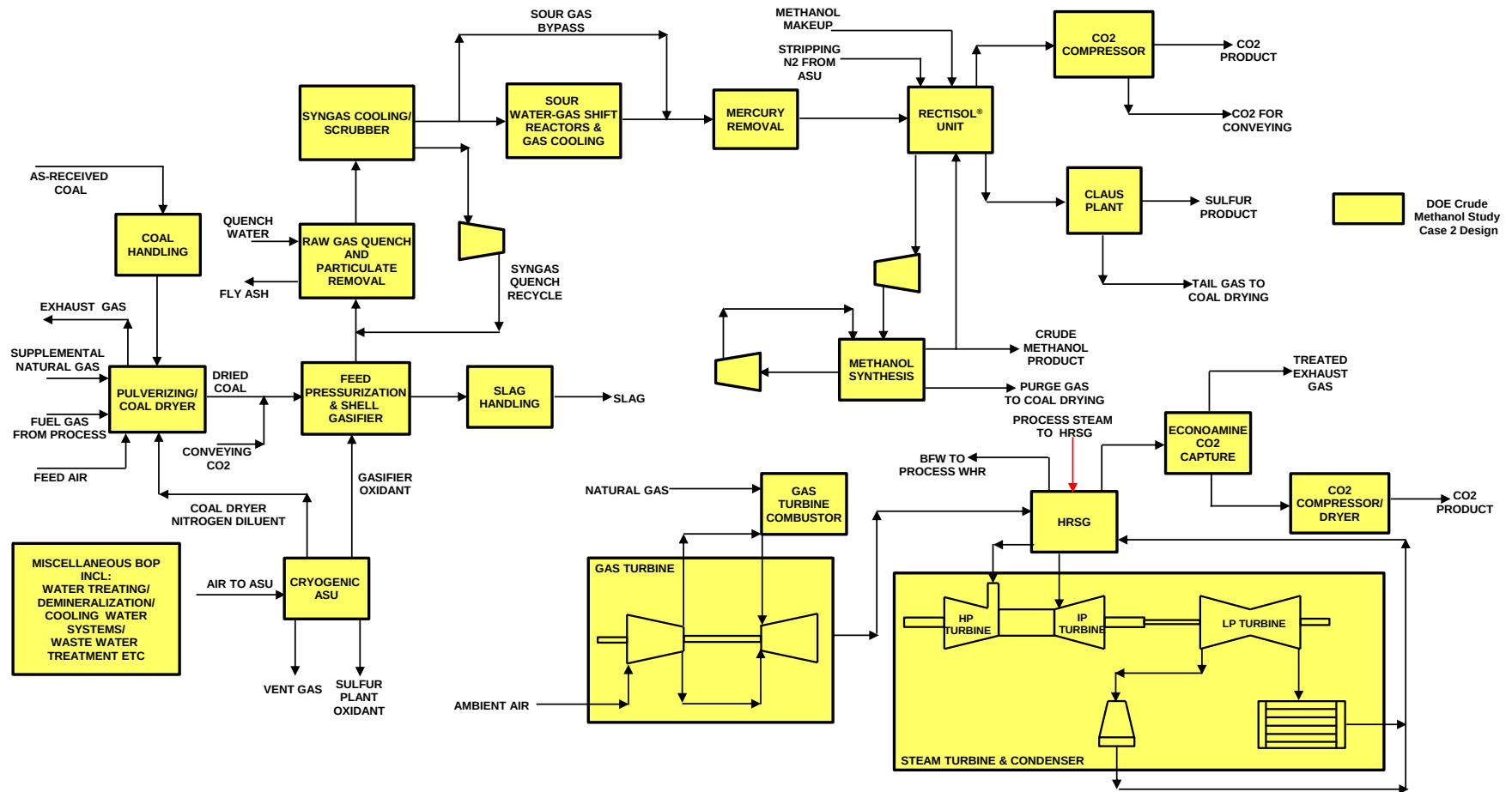
Table 2-2
Crude Methanol Study Case 2 Description

Case	Feedstock	Steam Cycle, psig/°F/°F	Combustion Turbine	Gasifier Technology	Oxidant	Sulfur Removal/ Recovery	CO ₂ Separation	Products
2	Coal/Syngas	1800/1050 /1050	SGT6-2000E	Shell	95mol% O ₂	Rectisol®/ Claus	Rectisol® & Amine ¹	Methanol

¹ Amine process is added to NGCC system only

A block flow diagram (BFD) of the Case 3a reference CTM plant is shown in Figure 2-1. It is assumed to operate with an annual on-stream factor of 90% or 7,884 hours/year at full capacity.

Figure 2-1
Case 3a: Reference Shell Gasifier with Rectisol® AGR CTM Plant - Simplified BFD



2.2.2 Case 3b: AR Compact Gasifier with RTI WDP CTM Plant

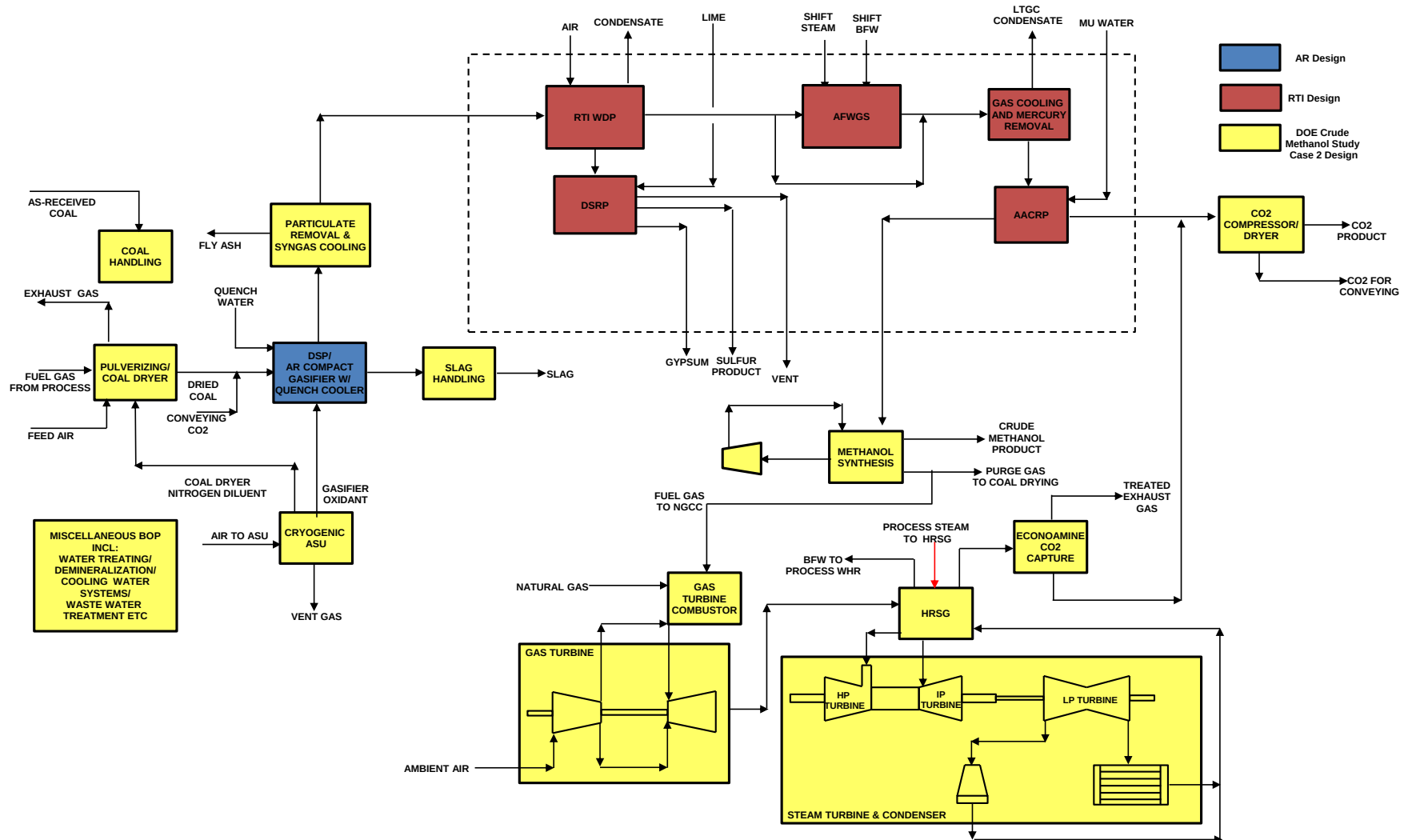
Case 3b is a preliminary conceptual design that fully integrates the AR gasification process with the RTI WDP. Due to the different cold gas efficiency of the AR gasifier, the CTM plant will consume a different amount of coal feed when compared with Case 3a in order to produce nominally, the same amount of syngas to produce 10,000 metric tons of methanol per day. For the NGCC plant, three General Electric (GE) MS6001B turbines were required to meet the plant's demands such that no power import is required. However, due to differences between the Case 3b plant's auxiliary power consumption, as well as differences in process waste heat recovery and associated steam generation from the Case 3a reference plant, some excess power is produced and exported to the grid for extra revenue.

A combination of RTI's WDP unit and AACRP unit replaces the Rectisol[®] unit in Case 3a to remove the sulfur and CO₂ from the syngas. The Case 3b plant will remove sulfur via RTI's WDP which removes hydrogen sulfide (H₂S) and carbonyl sulfide (COS) from the syngas after it leaves the particulate filters, without requiring additional cooling. The treated syngas undergoes sweet shift in a RTI's AFWGS process before it is cooled and sent to the AACRP unit for CO₂ capture. The AACRP will capture CO₂ equivalent to >90% of the raw syngas' carbon content in order to meet the CO₂ concentration requirements for methanol synthesis.

The Case 3b plant BFD is shown in Figure 2-2. This figure serves to demarcate the battery limits and highlights the interfaces between AR's and RTI's proprietary systems (colored blue and red respectively) and the rest of the CTM processes (in yellow) that are derived from the Case 3a reference plant (Case 2 of the Crude Methanol Study). The blue block represents AR's coal feed and compact gasifier system, which replaces the lockhopper feed system and Shell gasifier in the Case 3a reference plant. The red blocks within the broken-line rectangle represent RTI's advanced syngas cleanup and AACRP processes and comprise the WDP for sulfur removal, DSRP to produce elemental sulfur, high-temperature sweet WGS, low-temperature gas cooling (LTGC), and AACRP for CO₂ removal. The blue and red blocks in Figure 2-2 are the primary focus of AR and RTI's efforts, and they each provided the system designs for the units within these blocks in Figure 2-2.

The yellow blocks in Figure 2-2 are the balance of plant processes that are unrelated to the AR gasification and RTI advanced syngas cleanup systems. These blocks have the same unit operations as those in Case 3a, but with different throughputs due to the different flow quantities. Costs for these systems were scaled based on system capacities from Case 3a using factors given in the QGESS Capital Cost Scaling Methodology document wherever possible.

Figure 2-2
Case 3b: AR Compact Gasifier with RTI WDP CTM Plant - Simplified BFD



2.2.3 Case 3c: AR Compact Gasifier with Rectisol®-AGR CTM Plant

To complete the evaluation of benefits of the AR and RTI technologies, two other CTM cases were evaluated, each utilizing either the AR advanced gasification or RTI advanced syngas cleanup on a standalone basis. Case 3c is a design that uses the AR gasification process but paired with a two-stage Rectisol® process to remove both sulfur and CO₂ in the syngas. The coal feed rate is expected to be the same as Case 3b since both utilize the AR gasification process operating under the same conditions, hence same cold gas efficiency, to generate enough syngas for production of 10,000 tonnes per day of crude methanol. For the Case 3c NGCC plant, three GE MS6001B turbines were required to meet the plant's demands such that no power import is required, with excess power generated for sale.

The Case 3c plant BFD is shown in Figure 2-3. The blue block represents AR's DSP and compact gasifier system. AR provided the design and performance data for these systems. The unchanged yellow blocks consist of the balance of plant processes as in Case 3a but with different throughputs. Their costs are scaled based on system capacities using factors from the QGESS *Capital Cost Scaling Methodology* document.

2.2.4 Case 3d: Shell Gasifier with RTI WDP CTM Plant

Case 3d is the design that uses the Shell gasification process paired with RTI advanced syngas cleanup and AACRP processes to remove sulfur and CO₂ from the syngas. The coal feed rate is expected to be the same as Case 3a since both utilize Shell gasifiers, which, with the same cold gas efficiencies, generate similar syngas quantities for production of 10,000 tonnes per day of crude methanol in the methanol synthesis units. Three GE MS6001B gas turbines were required to meet the Case 3d plant's auxiliary demands such that no power import is required, with any excess power exported to the grid as byproduct to generate additional revenue.

The red blocks within the broken-line rectangle represent RTI's advanced syngas cleanup process and consist of the unit operations designed by RTI. These include the equipment for WDP, DSRP, AFWGS, LTGC and the AACRP unit. H₂S and COS are removed from the syngas in the WDP unit while primarily CO₂ is removed via the AACRP unit.

The unchanged yellow blocks consist of the balance of plant processes as in Case 3a but with different throughputs. Like the previous three cases, their costs will be scaled based on system capacities using factors from the QGESS *Capital Cost Scaling Methodology* document.

Figure 2-3
Case 3c: AR Compact Gasifier with Rectisol®-Based AGR CTM Plant - Simplified BFD

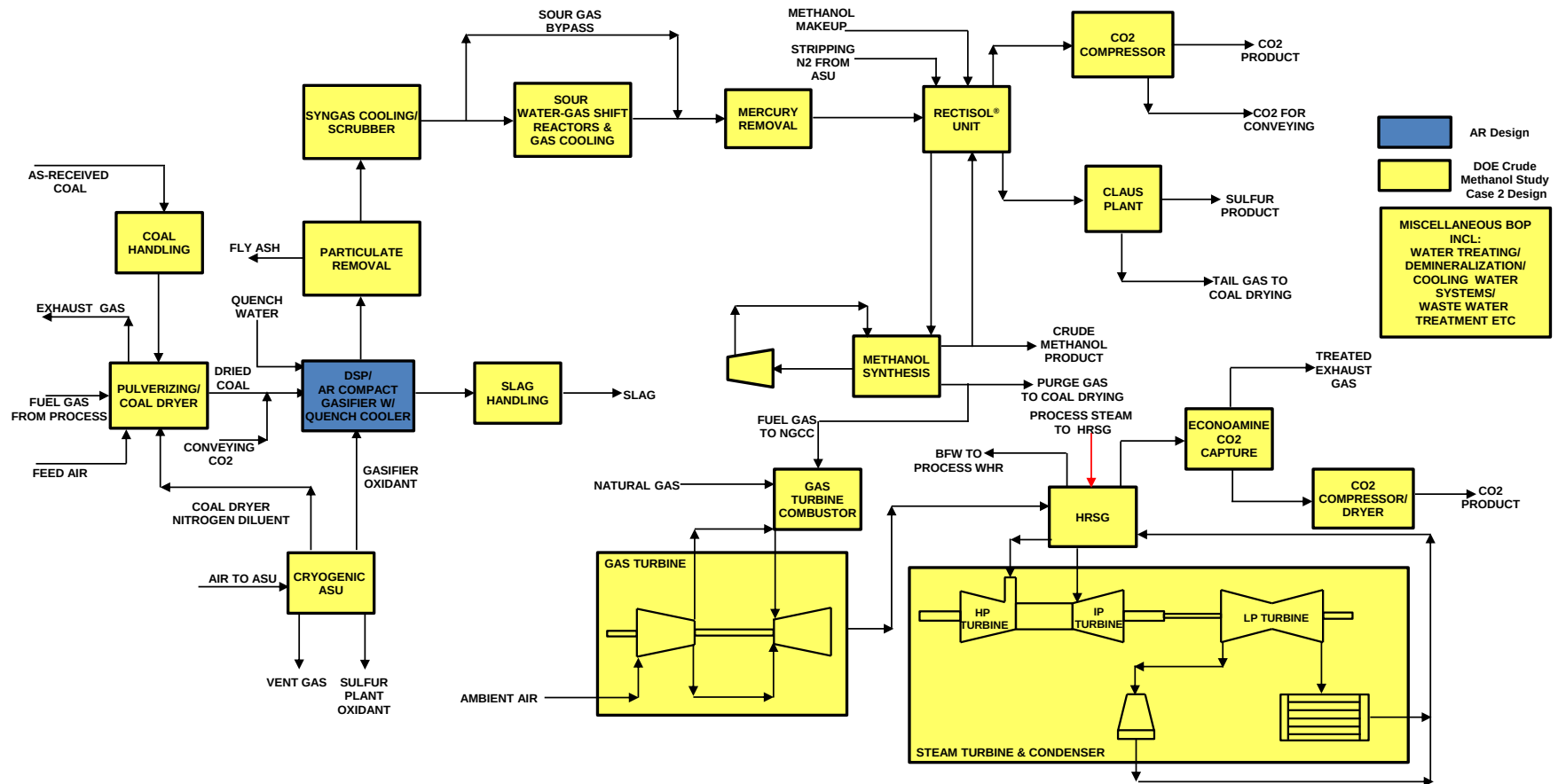
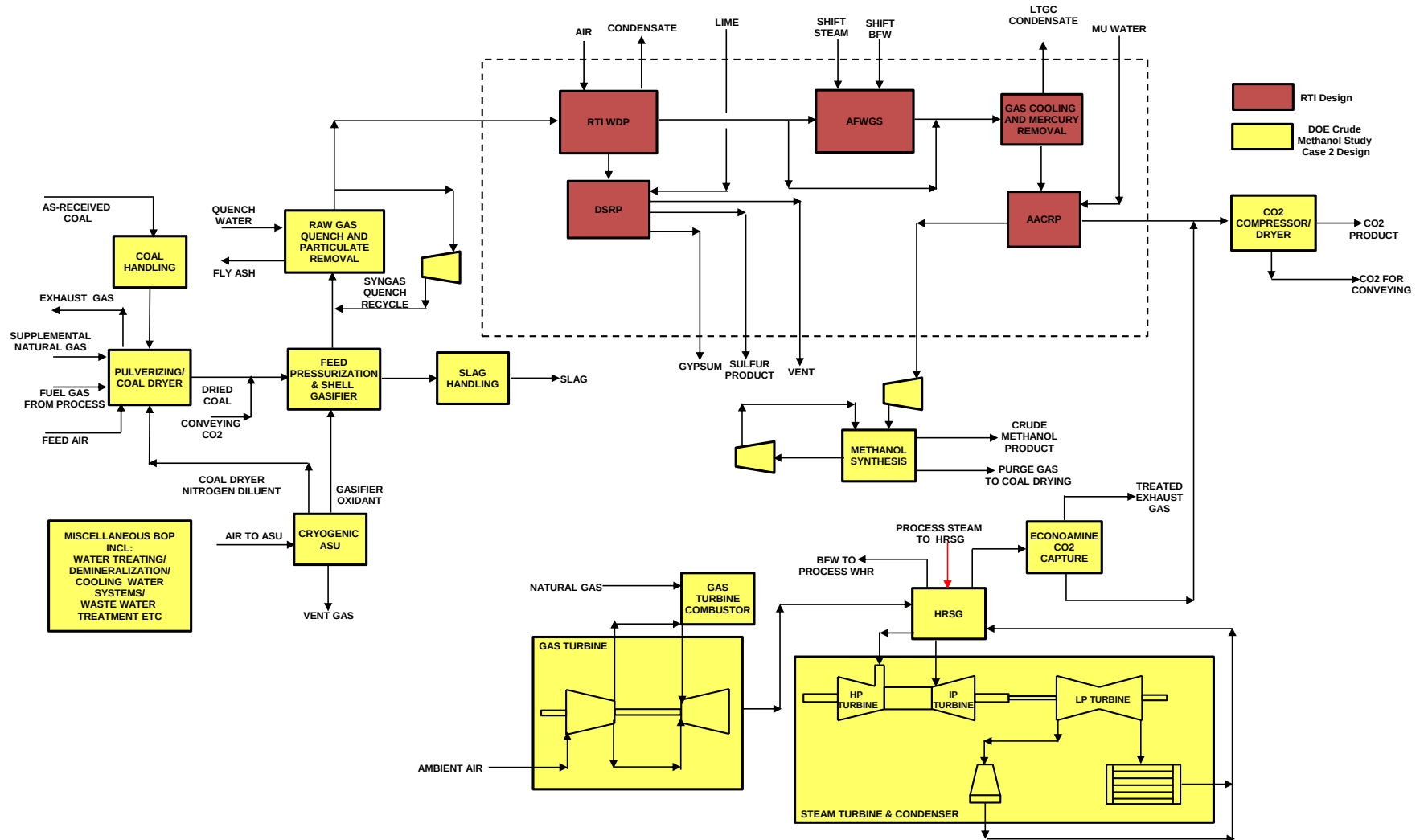


Figure 2-4
Case 3d: Shell Gasifier with RTI WDP CTM Plant - Simplified BFD



2.3 DESIGN CRITERIA

2.3.1 Heat and Material Balance

Nexant carried out simulation of all four CTM cases on ASPEN Plus to obtain the process heat and material balances (HMB). This helps to facilitate a more detailed estimation of the overall plant utility balance, including process waste heat recovery, generation from the power cycle, as well as cooling water load breakdown, all of which help to determine the overall CTM plant performance with more certainty.

Nexant developed a conceptual design of the Case 3a reference plant with two-stage Rectisol[®]-based AGR using data from Case 2 of the Crude Methanol Study as the basis. The resulting stream flows, HMB, methanol production and power generation from the NGCC were benchmarked and cross-checked against the Crude Methanol Study Case 2 to ensure that the results were within a reasonable range of accuracy.

From the benchmark design, Nexant provided AR and RTI with the stream conditions and flows to the gasification and WDP units respectively. Using these as inputs into their models, AR and RTI determined the heat and material balances around their respective process systems, overall utilities consumption, and equipment sizes. These outputs were then transmitted to Nexant, where these were plugged into its in-house CTM plant simulation to complete the modeling of Cases 3b, 3c, and 3d.

2.3.2 AR Compact Gasifier and Feed System

2.3.2.1 DSP Feed System

AR provided Nexant with the specifications and costs of its proprietary DSP. Nexant had previously completed a TEA (DE-F0012062) for DOE/NETL on behalf of AR that studied the performance and costs of the overall DSP feed system. In this study, Nexant integrated the AR DSP with the rest of the feed system equipment, including the atmospheric feed silo, pressurized feed silo, solid filters, etc. to arrive at the overall system cost and performance. For the current study, the performance, utilities consumption and costs for the AR DSP coal feed system were derived from that prior TEA.

2.3.2.2 AR Compact Gasification System

AR provided Nexant with their proprietary compact gasifier's input and output stream flows and conditions, enabling Nexant to integrate the gasifier into its model. AR also provided Nexant the gasifier's utilities consumption and equipment costs. These were used as inputs to help determine the overall CTM plant performance and total plant cost (TPC) for Cases 3b and 3c.

2.3.3 RTI WDP System

RTI provided Nexant with the major stream flows and conditions into, within and out of their proprietary advanced syngas cleanup process and AACRP, enabling Nexant to integrate these processes into its model. For cost estimations of RTI's advanced syngas cleanup and AACRP, RTI provided Nexant with their turnkey costs that were based on scale-up of actual cost data from their 50 MWe demonstration plant construction. RTI also provided Nexant with a list containing the advanced syngas cleanup and AACRP systems' utilities and consumables rates.

These were used as inputs to establish the overall CTM plant performance, TPC and variable operating costs for Cases 3b and 3d.

2.3.4 Other Systems

The costs for the balance of plant processes not directly related to the AR gasification, RTI advanced syngas cleanup, and AACRP systems were derived via capacity-factor from the corresponding system costs listed in the Crude Methanol Study and escalated to year 2011 costs.

2.4 SITE-RELATED CONDITIONS

The CTM plants in this study are assumed to be located in a generic, Midwestern USA site, with site-related conditions as shown below:

▪ Location	Greenfield, Midwestern USA
▪ Elevation, ft above sea level	0
▪ Topography	Level
▪ Size, acres	300
▪ Transportation	Rail
▪ Ash/slag disposal	Off Site
▪ Water	Municipal (50%)/Groundwater (50%)
▪ Access	Landlocked, having access by train and highway
▪ CO ₂ disposition	Compressed to 2,215 psia at battery limit before being transported 50 miles for sequestration in a saline formation at a depth of 4,055 ft. (Study scope limited to delivery at CTM plant battery limit only)

2.5 METEOROLOGICAL DATA

Maximum design ambient conditions for material balances, thermal efficiencies, system design and equipment sizing are:

▪ Barometric pressure, psia	14.696
▪ Dry bulb temperature (DBT)	59 °F
▪ Wet bulb temperature (WBT)	51.5 °F
▪ Ambient relative humidity, %	60

2.6 FUEL PROPERTIES

2.6.1 Coal Characteristics

Design coal feed to the CTM plants is Montana Powder River Basin (PRB) subbituminous coal with characteristics presented in Table 2-3. The as-received coal properties shown in Table 2-3 are from the QGESS Detailed Coal Specifications document. The as-received coal is dried to

6% moisture before being pressurized and fed to the gasifier. The gasifier will gasify enough dried PRB coal to produce sufficient syngas to produce 10,000 tonnes per day of methanol.

Table 2-3
Montana PRB Coal Specification

Rank	Subbituminous	
Seam	Montana Rosebud PRB	
Source	Western Energy Co.	
Ultimate Analysis, weight%	As-Received	Dried Coal to Gasifier
Carbon	50.07	63.40
Hydrogen	3.38	4.29
Nitrogen	0.71	0.90
Chlorine	0.01	0.01
Sulfur	0.73	0.92
Oxygen	11.14	14.11
Ash	8.19	10.37
Moisture	25.77	6.00
Total	100.0	100.0
Proximate Analysis, weight%	As-Received	Dried Coal to Gasifier
Volatile Matter	30.34	38.42
Fixed Carbon	35.70	45.20
Ash	8.19	10.38
Moisture	25.77	6.00
Total	100.0	100.0
Higher Heating Value (HHV), Btu/lb	8,564	10,825
Sulfur Analysis*, weight%		Dry
Pyritic		0.63
Sulfate		0.01
Organic		0.34
Mercury, ppmW (moisture-free basis)		0.081
Ash Fusion Temperatures at Reducing Conditions, °F		
Initial Deformation	2,238	
Softening	2,254	
Hemispherical	2,270	
Fluid	2,298	

*In accordance with DOE/NETL Crude Methanol Study, this study assumes that all sulfur in the coal is converted in the gasifier and leaves with the syngas

2.6.2 Natural Gas Characteristics

The natural gas composition used in this analysis, representative of natural gas after going through standard midstream processing, is presented in Table 2-4.

Table 2-4
Natural Gas Composition

Compositions:		Volume Percentage
Methane	CH ₄	93.1
Ethane	C ₂ H ₆	3.2
Propane	C ₃ H ₈	0.7
n-Butane	C ₄ H ₁₀	0.4
Carbon Dioxide	CO ₂	1.0
Nitrogen	N ₂	1.6
Total		100.0
LHV, Btu/lb		20,410
LHV, Btu/scf		932
HHV, Btu/lb		22,600
HHV, Btu/scf		1,032

2.7 CO₂ PRODUCT TREATING AND PURIFICATION DESIGN CRITERIA

For this study, recovered CO₂ is delivered at the battery limit (B/L), with specifications listed in Table 2-5 for saline reservoir sequestration per the QGESS CO₂ Impurities Design Parameters document.

Table 2-5
CO₂ Product Specifications

Compositions:		Conceptual Design	Range in Literature
CO ₂ ,	vol% (Min)	95	90 - 99.8
N ₂	vol% (Max)	} 4	0.01 - 7
Ar	vol% (Max)		0.01 - 4
O ₂ ,	vol% (Max)		0.01 - 4
H ₂	vol% (Max)		0.01 - 4
CH ₄ ,	vol% (Max)		0.01 - 4
H ₂ O,	ppmv (Max)	300	20 - 650
H ₂ S	vol% (Max)	0.01	0.002 - 1.3
NH ₃	ppmv (Max)	50	0 - 50
SO ₂ ,	ppmv (Max)	100	10 - 50,000
NO _x ,	ppmv (Max)	100	20 - 2,500
CO,	ppmv (Max)	35	10 - 5,000
B/L pressure, psia		2,215	

CO₂ compression and drying facilities were provided to boost the CO₂ product pressure to the required B/L requirement.

2.8 POWER GENERATION & AUXILIARY LOADS

For the power generation portion in each of the four cases in the current study, Nexant used its NGCC model, which takes into account the power requirement of the entire plant, process steam generated in the CTM island, as well as steam consumptions in both the CTM and power islands. Auxiliary loads were estimated, wherever applicable, by pro-rating from the Case 2 of the Crude Methanol Study using relevant scaling parameters obtained from the model's HMB.

Table 2-6 shows the power production and auxiliary load breakdown of the original Case 2 of the Crude Methanol Study, which Case 3a of this study was modeled upon and benchmarked against. For reference purposes, the scaling parameters are also shown in Table 2-6.

The AR compact gasification system and RTI advanced syngas cleanup and AACRP systems have different auxiliary loads that are not scalable with some of the loads specified in the Case 2 of the Crude Methanol Study. AR and RTI provided the auxiliary loads for their systems and these were used directly as inputs to the auxiliary load calculation.

Table 2-6
Crude Methanol Study Case 2 Auxiliary Load and Power Production Summary

Power Cycle Power Generation	Electrical Load, kWe	Scaling Capacity
Gas Turbine Power	113,700	<i>Calculated from Model</i>
Steam Turbine Power	276,400	<i>Calculated from Model</i>
Total	390,100	
Auxiliary Load Description	Electrical Load, kWe	Scaling Capacity
CTM AUXILIARY LOADS		
Coal Handling & Milling	9,090	As Received Coal
Slag Handling	1,940	Slag Flow
Air Separation Unit	179,940	Oxygen Production
Syngas Recycle Compressor	6,600	Recycle Syngas Flow
Incinerator Air Blower	2,680	Incinerator Air Flow
Direct-Fired Boiler Air Blower	310	Direct Fired Boiler Air Flow
Flash Bottoms Pump	720	Flash Bottoms Flow
Scrubber Pumps	1,070	Scrubber Flow
Rectisol® Auxiliary	51,270	<i>Calculated from Vendor Data</i>
Claus Plant Auxiliary	250	Sulfur Flow
CO ₂ Compressor Auxiliary	68,820	CTM CO ₂ Product Flow
Syngas Compressor	20,760	Syngas Flow to Methanol Unit
Recycle Gas Compressor	3,370	Recycle Gas Flow
Water Treatment	3,530	Total Wastewater + MU Water Flow
Air Cooler Fans	1,800	Gas Cooling Duty
Circulating Water Pump	9,430	CTM Circulating Water Flow
Boiler Feed Water Pump	1,500	CTM BFW Flow
Cooling Tower Fans	510	CTM Cooling Duty
Steam Turbine Auxiliaries	100	CTM Steam Turbine Production
Miscellaneous BOP	5,000	Fixed
SUBTOTAL CTM AUXILIARIES, kWe	368,690	
NGCC AUXILIARY LOADS		
Condensate Pumps	210	<i>Calculated from Model</i>
Boiler Feedwater Pumps	3,910	<i>Calculated from Model</i>
Amine System Auxiliaries	3,000	CO ₂ Product Flow
NGCC CO ₂ Compression	4,800	CO ₂ Product Flow
Circulating Water Pump	3,730	<i>Calculated from Model</i>
Ground Water Pumps	350	Ground Water Flow
Cooling Tower Fans	2,030	Power Cycle Cooling Duty
SCR	10	Fixed
Gas Turbine Auxiliaries	700	Gas Turbine Output
Steam Turbine Auxiliaries	100	Steam Turbine Output
Miscellaneous BOP	500	Fixed
Transformer Losses	2,140	Power Cycle Gross Output
SUBTOTAL NGCC AUXILIARIES, kWe	21,480	
TOTAL AUXILIARY POWER, kWe	390,170	
NET PLANT POWER, kWe	-70	

2.9 RAW WATER SUPPLY

Raw water makeup was assumed to be provided 50% by a publicly owned treatment works and 50% from groundwater.

2.10 ENVIRONMENTAL DESIGN TARGETS

The environmental limits presented in this section refer to the gasification/power cycle only, because the environmental requirements for the methanol plant are considered beyond the scope of the study.

The environmental targets for the study were considered on a technology- and fuel-specific basis. Since all the cases are located at a greenfield site, permitting a new plant would involve the New Source Review (NSR) permitting process. The NSR process requires installation of emission control technology, meeting either the best available control technology (BACT) determinations for new sources located in areas meeting ambient air quality standards, or the lowest achievable emission rate (LAER) technology for sources located in areas that do not meet ambient air quality standards. This study uses the BACT guidelines, summarized in Table 2-7.

Table 2-7
BACT Environmental Design Basis

Pollutant	Control Technology	Limit
Sulfur Oxides (SO ₂)	Rectisol® + Claus Plant/ Econamine Plus	≤ 0.050 lb/MMBtu
Nitrogen Oxides (NO _x)	Low NO _x Burner (LNB) and N ₂ dilution primarily with humidification as needed	15 ppmbd (at 15% O ₂)
Particulate Matter (PM)	Cyclone/Barrier Filter/Wet Scrubber/AGR Absorber	0.006 lb/MMBtu
Mercury (Hg)	Activated Carbon Bed	95% removal

Total air pollutants in all vents must meet the above specifications even if atmospheric venting is minimal.

2.11 OTHER SITE SPECIFIC REQUIREMENTS

Although the following design parameters are considered site-specific, and are not quantified for this study, allowances for normal conditions and construction are included in the cost estimates.

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Fire protection
- Local code height requirements
- Noise-regulations – Impact on site and surrounding area

2.12 CAPITAL COST ESTIMATION METHODOLOGY

2.12.1 General

For the CTM plant with CO₂ capture, the Crude Methanol Study provided a code of accounts grouped into 14 major systems. Each of these major systems is broken down further into different subsystems. This type of code of accounts structure has the advantage of grouping all reasonably allocable components of a system or process into a specific system account.

For the four CTM cases evaluated in this study, capital cost scaling following the guidelines and parameters that are described in the NETL Capital Cost Scaling Methodology document was used to perform the cost estimation for systems that are not related to the AR advanced gasification and RTI advanced syngas cleanup and AACRP systems. In general, this cost estimation methodology involves determining the scaling parameters, exponents and coefficients from the Capital Cost Scaling Methodology, as well as the reference cost and baseline capacity from the Crude Methanol Study. Once these have been established, the capital cost can be estimated based on the revised capacity from the HMB developed by Nexant's ASPEN models of the CTM cases.

Nexant performed a bottoms-up, major-equipment factored cost estimation for the AR gasifier and RTI advanced syngas cleanup and AACRP systems. These are described in greater detail in Section 2.12.2 and 2.12.3 respectively.

2.12.2 AR Gasification System Capital Cost Estimate Criteria

2.12.2.1 AR-Licensed Systems Cost Estimates

Costs for equipment in the AR advanced gasification system that are proprietary to AR, such as its DSP and compact gasifier, were provided by AR. Nexant used and reported these costs on an as-provided basis. The level of detail provided in these cost estimates was determined by AR itself.

2.12.2.2 Nexant-Designed System Cost Estimates

For auxiliary equipment within the AR gasification section that is not included in the technology licensor's package, Nexant performed a bottom-up cost estimate based on equipment sizing. Based on the equipment sizes defined by process HMB, cost for each piece of major equipment was estimated based on either vendor quotes or using commercial estimation software (ASPEN In-Plant Cost Estimator). Installation labor hours for the equipment were factored from historical data by equipment type.

An average labor wage at \$39.7/hr, with an all-in labor cost of \$51.6/hr (including wages plus 30% burden to cover fringe benefits, payroll based taxes, and insurance premiums) was assumed for calculating the 2011 installation labor costs. No over-time or other premiums were added. The average labor productivity for the site was assumed to be 105% of the US Gulf coast productivity.

Bulk material and installation costs were factored from major equipment cost (MEC). Bulk materials that cover instrumentation, piping, structure steel, insulation, electrical, painting, and concrete and site preparation work needed to complete the major equipment installations were

factored from MEC based on historical data for similar services. Installation labor for each bulk commodity was factored from historical data by type. Sum total of MEC plus bulk material costs plus installation labor costs forms the total direct cost (TDC) for the feed system.

Construction indirect costs were then factored from total direct labor costs based on historical data, and added to the system TDC to give the total field cost (TFC) for the system. Construction indirect cost covers the cost for setup, maintenance and removal of temporary facilities, warehousing, surveying and security services, maintenance of construction tools and equipment, consumables and utilities purchases, and field office payrolls. It should be noted that the term TFC is the equivalent of the Bare Erected Cost (BEC) used in the Crude Methanol Study.

2.12.3 RTI WDP System Capital Cost Estimate Criteria

For WDP, RTI's process model incorporates actual data from the pilot plant testing at Eastman Chemical and the 50 MWe demonstration plant at Tampa Electric Company. For AACRP, RTI developed a detailed process model using ProMax with TSWEET kinetics for the CO₂-amine reactions. This process model was reviewed by a recognized expert in the field of acid gas removal technologies. RTI has diligently worked to validate these models against actual pilot and demonstration plant data for RTI's proprietary technology and any publicly available DOE/NETL reports for the more commercial technologies.

Using these models, RTI developed HMBs and sized equipment lists. Using process conditions and experience gained from the pilot plant and demonstration plant operation, materials of construction were identified. With the sized equipment list and materials of construction, RTI developed equipment cost estimates from actual vendor quotes or scaled equipment costs from the 50 MWe demonstration unit for a majority of the equipment. For some of the more generic and non-proprietary equipment, the equipment costs were developed using Aspen Plus ICARUS Economic Analyzer. RTI then used the same factored estimation methods used in the Crude Methanol Study to estimate BEC for WDP, AFWGS, and AACRP. This BEC information was used by Nexant directly in its calculation of the overall CTM plant capital expenditure (CAPEX).

In addition to using the actual pilot plant and demonstration plant data for developing equipment costs, RTI used this available data to develop utility and variable cost especially for sorbent, catalysts, and solvent. Estimates of the utilities and variable costs were also provided to Nexant. Nexant used these to complete additional performance analysis and calculate the CTM plant's overall variable operating costs.

2.12.4 Balance of Plant Capital Cost Estimate Criteria

The capital cost estimates for the rest of the CTM systems that are unrelated to coal gasification, sulfur and CO₂ removal, were developed based on the Case 2 Shell CTM plant with CO₂ capture case in the Crude Methanol Study. The costs were adjusted for differences in unit or plant capacity according to NETL's Guidelines as described in the NETL Capital Cost Scaling Methodology document.

Table 2-8 shows the code of accounts for the CTM plant. These systems are further broken down to include the various subsystems. The scaling parameters for these balance-of-plant (BOP)

processes, as laid out by the NETL Capital Cost Scaling Methodology document, are also shown in Table 2-8.

Table 2-8
Code of Accounts for Report CTM Plant

Acct No.	Item/Description	Scaling Parameter
1	COAL & SORBENT HANDLING	
1.1	Coal Receive & Unload	Coal Feed Rate
1.2	Coal Stackout & Reclaim	Coal Feed Rate
1.3	Coal Conveyors & Yard Crush	Coal Feed Rate
1.4	Other Coal Handling	Coal Feed Rate
1.9	Coal & Sorbent Handling Foundations	Coal Feed Rate
2	COAL & SORBENT PREP & FEED	
2.1	Coal Crushing & Drying	Coal Feed Rate
2.2	Prepared Coal Storage & Feed	Coal Feed Rate
2.3	Dry Coal Injection System	Calculated
2.4	Misc Coal Prep & Feed	Coal Feed Rate
2.9	Coal & Sorbent Feed Foundation	Coal Feed Rate
3	FEEDWATER & MISC. BOP SYSTEMS	
3.1	Feedwater System	BFW (HP only)
3.2	Water Makeup & Pretreating	Raw Water Makeup
3.3	Other Feedwater Subsystems	BFW (HP only)
3.4	Service Water Systems	Raw Water Makeup
3.5	Other Boiler Plant Systems	Raw Water Makeup
3.6	FO Supply Sys and Nat Gas, incl Pipeline	Coal Feed Rate
3.7	Waste Treatment Equipment	Raw Water Makeup
3.8	Misc Power Plant Equipment	Gross Power Output
4	GASIFIER & ACCESSORIES	
4.1	Gasifier, Syngas Cooler & Auxiliaries	Syngas Throughput
4.3	ASU/Oxidant Compression	O ₂ Production
4.4	LT Heat Recovery & FG Saturation	Syngas Flow
4.6	Flare Stack System	Syngas Flow
4.9	Gasification Foundations	Syngas Flow
5A	GAS CLEANUP & PIPING	
5A.1	Rectisol® System	Gas Flow to AGR
5A.2	Elemental Sulfur Plant	Sulfur Production
5A.3	Mercury Removal	Hg Bed Carbon Fill
5A.4a	COS Hydrolysis	COS Catalyst
5A.4b	Shift Reactors	WGS Catalyst
5A.5	Pressure Swing Adsorption System	Syngas Flow
5A.6	Blowback Gas System	Syngas Flow
5A.7	Fuel Gas Piping	Fuel Gas Flow
5A.9	HGCU Foundations	Sulfur Production
5B	CO ₂ REMOVAL & COMPRESSION	
5B.1	NGCC CO ₂ Removal System	NGCC CO ₂ Flow
5B.2	CO ₂ Compression & Drying	Total CO ₂ Product
5C	METHANOL PRODUCTION	
5C.1	Methanol Synthesis	Methanol Product

Acct No.	Item/Description	Scaling Parameter
6	COMBUSTION TURBINE/ACCESSORIES	
6.1	Combustion Turbine Generator	Natural Gas Flow
6.9	Combustion Turbine Foundations	Natural Gas Flow
7	HRSG, DUCTING & STACK	
7.1	Heat Recovery Steam Generator	HRSG Duty
7.2	HRSG Accessories	HRSG Duty
7.9	HRSG, Duct & Stack Foundations	Vol Flow to Stack
8	STEAM TURBINE GENERATOR	
8.1	Steam TG & Accessories	Turbine Capacity
8.2	Turbine Plant Auxiliaries	Turbine Capacity
8.3a	Condenser & Auxiliaries	Condenser Duty
8.3b	Air Cooled Condenser	Condenser Duty
8.4	Steam Piping	BFW (HP Only)
8.9	TG Foundations	Turbine Capacity
9	COOLING WATER SYSTEM	
9.1	Cooling Towers	Cooling Tower Duty
9.2	Circulating Water Pumps	Circ H ₂ O Flow Rate
9.3	Circ. Water System Auxiliaries	Circ H ₂ O Flow Rate
9.4	Circ. Water Piping	Circ H ₂ O Flow Rate
9.5	Makeup Water System	Raw Water Makeup
9.6	Component Cooling Water System	Circ H ₂ O Flow Rate
9.9	Circ. Water System Foundations	Circ H ₂ O Flow Rate
10	ASH/SPENT SORBENT HANDLING SYS	
10.1	Slag Dewatering & Cooling	Slag Production
10.6	Ash Storage Silos	Slag Production
10.7	Ash Transport & Feed Equipment	Slag Production
10.8	Misc. Ash Handling System	Slag Production
10.9	Ash/Spent Sorbent Foundation	Slag Production
11	ACCESSORY ELECTRIC PLANT	
11.1	Generator Equipment	Turbine Capacity
11.2	Station Service Equipment	Auxiliary Load
11.3	Switchgear & Motor Control	Auxiliary Load
11.4	Conduit & Cable Tray	Auxiliary Load
11.5	Wire & Cable	Auxiliary Load
11.6	Protective Equipment	Auxiliary Load
11.7	Standby Equipment	Total Gross Output
11.8	Main Power Transformers	Total Gross Output
11.9	Electrical Foundations	Total Gross Output
12	INSTRUMENTATION & CONTROL	
12.4	Other Major Component Control	Auxiliary Load
12.6	Control Boards, Panels & Racks	Auxiliary Load
12.7	Computer & Accessories	Auxiliary Load
12.8	Instrument Wiring & Tubing	Auxiliary Load
12.9	Other I & C Equipment	Auxiliary Load
13	IMPROVEMENT TO SITE	
13.1	Site Preparation	Accounts 1-12
13.2	Site Improvements	Accounts 1-12
13.3	Site Facilities	Accounts 1-12

Acct No.	Item/Description	Scaling Parameter
14	BUILDING & STRUCTURES	
14.1	Combustion Turbine Area	Gas Turbine Power
14.2	Steam Turbine Building	Accounts 1-12
14.3	Administration Building	Accounts 1-12
14.4	Circulation Water Pump House	Circ H ₂ O Flow Rate
14.5	Water Treatment Buildings	Raw Water Makeup
14.6	Machine Shop	Accounts 1-12
14.7	Warehouse	Accounts 1-12
14.8	Other Buildings & Structures	Accounts 1-12
14.9	Waste Treating Building & Structures	Raw Water Makeup

As Table 2-8 is based on Case 2 of the Crude Methanol Study, it does not necessarily have account and/or subaccount numbers for the advanced technologies being evaluated in this study. To support optimal comparison of these advanced technologies with the existing commercial technologies, the advanced technologies used the same account and/or subaccount numbers as the existing commercial technologies that they are most analogous to. For technologies without a defined account number, one was created. Table 2-9 provides a list of the advanced technologies evaluated in this study and their associated account numbers.

Table 2-9
Code of Accounts for Advanced Technologies being Evaluated

Account Number	Title	Advanced Technology
2.3	Dry Coal Injection System	AR DSP
4.1	Gasifier, Syngas Cooler & Auxiliaries	AR Compact Gasifier
5A.1	AACRP	AACRP
5A.2	RTI DSRP	RTI DSRP
5A.4a	RTI WDP System	RTI WDP
5A.4b	Shift Reactors	AFWGS

2.12.5 Home Office, Engineering Fees and Project/Process Contingencies

Engineering and construction management fees and home office cost, project and process contingencies were factored from each subsystem's TFC. These were then added to the TFC to come up with the total plant cost (TPC) of the system.

Process contingency is intended to compensate for uncertainty in cost estimates caused by performance uncertainties associated with the development status of a technology. Process contingency for the advanced technologies of interest in this study were developed according to the criteria recommended by the Association for Advancement of Cost Engineering International (AACE) as specified in DOE/NETL's *Cost Estimation Methodology for NETL Assessment of Power Plant Performance*. Table 2-10 shows the process contingencies applied for the advanced technologies in this study.

Table 2-10
Process Contingencies for Advanced Technologies being Evaluated

Account Number	Advanced Technology	Process Contingency (% of Associated Process Capital)	Rationale
2.3	AR DSP	20%	The AR DSP and static splitter feed system are currently demonstrated at the small pilot level. The system shares similar types of commercially demonstrated equipment with the conventional lockhopper feed system e.g feed bins, compressors.
4.1	AR Compact Gasifier	25%	At 25%, this is within range of contingencies applied to technologies demonstrated at the small pilot level
5A.1	AACRP	20%	This is consistent with the contingency applied for similar processes such as Selexol™ and Rectisol® in the DOE/NETL studies.
5A.2	RTI DSRP	20%	The DSRP technology has been demonstrated in RTI's pilot testing at Eastman Chemical Company (DE-AC26-99FT40675)
5A.4a	RTI WDP	20%	The WDP technology has been demonstrated in RTI's 50 MWe demonstration plant project (DE-FE0000489)
5A.4b	AFWGS	0%	Fixed-bed sweet water gas shift technology is commercially proven

2.12.6 Owner's Cost

Owner's cost was then added to TPC to come up with the total overnight cost (TOC) for the system. Owner's costs as defined in the Crude Methanol Study include the following:

- Preproduction costs –
 - 6 months of all labor cost
 - 1 month of maintenance materials
 - 1 month of non-fuel consumables
 - 1 month of waste disposal
 - 25% of 1 month fuel cost at 100% capacity factor (CF)
 - 2% TPC
- Inventory capital -
 - 60 day supply of fuel and consumable at 100% CF
 - 0.5% TPC (spare parts)
- Initial cost for catalyst, sorbent and chemicals per design
- Land cost = \$900,000 (300 acres x \$3,000/acre)
- Other owner's costs at 15% TPC
- Financing costs at 2.7% TPC

2.13 OPERATION & MAINTENANCE COSTS

The operation and maintenance (O&M) costs pertain to those charges associated with operating and maintaining the CTM plants over their expected life. These costs include:

- Operating labor
- Maintenance – material and labor
- Administrative and support labor
- Consumables
- Fuel
- Waste disposal

There are two components of O&M costs; fixed O&M, which is independent of methanol and power generation, and variable O&M, which is proportional to methanol and power generation. Variable O&M costs were estimated based on 90% CF.

2.13.1 Fixed Costs

Operating labor cost was determined based on the number of operators required to work in the plant. Other assumptions used in calculating the total fixed cost include:

• 2011 Base hourly labor rate, \$/hr	\$39.7
• Length of work-week, hr	50
• Labor burden, %	30
• Administrative/Support labor, % O&M Labor	25
• Maintenance material + labor, % TPC	2.4
• Maintenance labor only, % maintenance material + labor	40
• Property Taxes and insurances, % TPC	2

2.13.2 Variable Costs

The cost of consumables, including fuel, was determined based on the individual rates of consumption, the unit cost of each specific consumable commodity, and the plant annual operating hours. Waste quantities and disposal costs were evaluated similarly to the consumables.

The unit costs for major consumables and waste disposal was selected from the Crude Methanol Study Case 2, QGESS Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases and from the QGESS Fuel Prices for Selected Feedstocks in NETL Studies document.

The 2011 coal price as delivered to the Midwestern, USA CTM plant is \$36.57/ton, per the QGESS Fuel Prices for Selected Feedstocks in NETL Studies document. The 2011 natural gas price delivered to the same plant is \$6.13/MMBtu (Higher Heating Value [HHV]).

2.13.3 CO₂ Transport and Storage Costs

CO₂ Transport, Storage, and Monitoring (TS&M) costs used for the Midwestern plant location is \$11/tonne. The product methanol required selling prices are reported both with and without the cost of CO₂ TS&M.

2.14 FINANCIAL MODELING BASIS

2.14.1 Required Selling Price

The figure-of-merit used to evaluate the CTM plant's overall financial performance is the methanol's required selling price (RSP), expressed in \$/gal of crude methanol. All costs are expressed in the "first-year-of-construction" year dollars, and the resulting RSP is also expressed in "first-year-of-construction" year (2011) dollars.

The RSP was calculated for each case assuming (i) a financial structure representative of a commercial fuels project, and (ii) a financial structure with loan guarantees or other government subsidies. The financial assumptions and structures used to estimate the RSPs are shown in Table 2-11.

Table 2-11
Financial Assumptions for RSP Calculation

Parameter	Value
TAXES	
Income Tax Rate	38% (Effective, 34% Federal, 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	None
Tax Holiday	None
FINANCING TERMS	
Repayment Term of Debt	30 years
Grace Period on Debt Repayment	None
Debt Reserve Fund	None
TREATMENT OF CAPITAL COSTS	
Capital Cost Escalation During Construction (nominal annual rate)	3.6%
Distribution of Total Overnight Capital over the Capital Expenditure Period (before escalation)	5 Year Period: 10%, 30%, 25%, 20%, 15%
Working Capital	Zero for all parameters
% of Total Overnight Capital that is Depreciated	100%
INFLATION	
RSP, O&M, Fuel Escalation (nominal annual rate)	3.0% RSP, O&M, COE, Fuel
FINANCIAL STRUCTURE (COMMERCIAL FUELS PROJECTS)	
Type of Security	Percent of Total
Debt	50
Equity	50
FINANCIAL STRUCTURE (LOAN GUARANTEE PROJECTS)	
Type of Security	Percent of Total
Debt	60
Equity	40

The same financial modeling methodology was used for this study as per the Crude Methanol Study. This is a simplified method that is a function of the plant TPC, capital charge factor (CCF), fixed and variable operating costs, CF, and net power generation, as shown in the equation below:

$$RSP = \frac{\text{first year capital charge} + \text{first year fixed operating costs} + \text{first year variable operating costs}}{\text{annual gallons of crude methanol produced}}$$

$$RSP = \frac{(CCF)(TOC) + OC_{fix} + (CF)(OC_{var})}{(CF)(gal/yr MeOH)}$$

The CCF used in evaluating the RSPs were pre-calculated using the NETL Power Systems Financial Model (PSFM). These factors are valid for global economic assumptions used for the

two pre-determined finance structures and capital expenditure periods. The resulting CCFs for the commercial fuels and loan guarantee finance structures are 0.218 and 0.170, respectively.

2.14.2 CO₂ Sales Price

Sensitivity analysis was to be done to determine the impact of CO₂ sales on CTM plant's RSP. The varying parameter is the CO₂ sales price at the CTM plant gate and ranged between \$0/tonne (baseline case assuming no value to the product CO₂) and \$60/tonne.

The formula used to calculate the revised RSP after taking into account CO₂ sales is shown below:

$$RSP_{CO_2 \text{ Sales}} = \text{Baseline RSP} - \frac{(\text{CO}_2 \text{ Sales Price}) \times \text{annual tonnes of CO}_2 \text{ product}}{\text{annual gallons of crude methanol produced}}$$

2.14.3 Cost of CO₂ Emissions

Sensitivity analysis on cost of CO₂ emissions was also performed. The varying parameter is the CO₂ emissions cost. The range of the emissions cost is between \$0/tonne (baseline case assuming no CO₂ emissions cost) and \$60/tonne.

The formula used to calculate the revised RSP after taking into account the cost of CO₂ emissions is shown below:

$$RSP_{CO_2 \text{ Emissions}} = \text{Baseline RSP} + \frac{(\text{Cost of CO}_2 \text{ Emissions}) \times \text{annual tonnes of CO}_2 \text{ emitted}}{\text{annual gallons of crude methanol produced}}$$

2.14.1 Cost of CO₂ Captured/Avoided

As the CO₂ sales price increases, the methanol RSP values decrease and approach the without-capture RSP values. The cost of CO₂ captured can be interpreted as the breakeven plant gate sale price where the cost of capture equals the revenue generated by selling the recovered CO₂.

The cost of CO₂ captured is defined as the plant gate CO₂ sale price where each capture case equals its corresponding without-capture RSP, **excluding CO₂ TS&M costs**. The cost of CO₂ avoided has the same definition, except that it **includes CO₂ TS&M costs**. Based on this definition, the cost of CO₂ avoided is always greater than the cost of CO₂ captured.

For the non-capture cases, the size of the NGCC and its fuel consumption should be reduced to match the lower parasitic power demand from elimination of the NGCC CO₂ capture and CO₂ compression processes. However, as the task of modeling four corresponding non-capture CTM plants was outside of this study's scope, a simplified calculation was used to estimate each non-capture plant's methanol product RSP.

In this calculation, it was assumed that while each non-capture case eliminated the CO₂ compression and NGCC post-combustion CO₂ capture (PCC) systems, the size of its NGCC and resulting natural gas consumption remains the same as the corresponding case with CO₂ capture.

The resulting non-capture CTM plant still uses the same amount of coal and natural gas fuel but directly vents the NGCC flue gas and captured CO₂ from the process AGR unit. The capital costs associated with CO₂ compression and the PCC processes are also eliminated, reducing the overall plant CAPEX. Additionally, the auxiliary power demands of these systems are eliminated, resulting in a CTM plant with more power export to the grid. Both these factors result in a lower methanol RSP for the non-capture CTM plant.

It should be noted that an alternative to the method described above would be to use over-the-fence costs for NGCC-generated power with and without CO₂ capture. If DOE had provided either of these costs, this would simplify the calculations for both the RSP and cost of CO₂ capture as it would effectively eliminate the NGCC modeling requirement.

3.1 PROCESS OVERVIEW

The conversion of coal-to-methanol is a two-step process: first conversion of the coal to the appropriate quality syngas via gasification and applying the water gas shift (WGS) reaction, and second, converting the syngas to methanol by catalytic conversion.

Syngas is generated in the Case 3a reference plant from the gasification of PRB coal in a high-pressure, oxygen-blown Shell quench-type gasifier. The high temperature entrained-bed gasifier uses a partial water quench and syngas cooler to cool the hot syngas stream and generate steam for the WGS reactors and power generation. Crude raw syngas (post-quench) from the gasification unit is scrubbed and split into two streams. The first stream is fed to a sour water-gas shift reactor to increase the hydrogen content so that a hydrogen/carbon monoxide (H_2/CO) molar ratio of 2:1 in the feed stream to the methanol synthesis reactor can be achieved, while the second stream bypasses the shift reactors. The streams are combined downstream to achieve the desired composition. The partial bypass mode of operation allows the shift reactor to operate at a higher conversion ratio resulting in a smaller size. The syngas is cooled in a low-temperature heat recovery system and then cleaned of mercury in a fixed-bed reactor. Sulfur and CO_2 are removed from the syngas via the two-stage Rectisol® process in preparation for methanol synthesis. The treated syngas is fed into fixed-bed methanol synthesis reactors to generate methanol.

As with Case 2 in the Crude Methanol Study, the Case 3a reference plant uses an NGCC to generate sufficient power to meet the auxiliary loads and make the plant power neutral. Consequently, the Case 3a reference plant uses both coal and natural gas in the production of methanol and has the necessary processes to capture more than 90% of the carbon in the coal and natural gas.

The Case 3a reference plant is assumed to operate with an annual on-stream CF of 90% or 7,884 hr/year at full capacity.

3.2 CTM PLANT COMMON PROCESS AREAS

As shown in Figure 2-1 in Section 2.2.1, the Case 3a reference plant consists of the following major process and/or utility blocks. Some of these blocks, or process areas, are common to the Case 3b, 3c and 3d plant configurations. These common process areas are in bold and italicized.

- Coal Sizing Handling
- Coal Prep, Drying
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- Dry Coal Feed & Shell SCGP Gasifier System
- Syngas Cooling (Quench, Scrubbing, Steam Generation)
- Gas Cleaning (Filters, WGS & AGR)

-
- Mercury Removal
 - CO₂ Compression and Purification Facilities
 - Sulfur Plant
 - Methanol Reactor and Synthesis Loop
 - Natural Gas Combined Cycle (NGCC)
 - HRSG, Ducting and Stack
 - Cooling Water Systems
 - BFW/Condensate System
 - Slag Recovery and Handling
 - Accessory Electric Plant
 - Instrumentation and Control

The common areas are presented in brief here for general background information, and to avoid unnecessary repetition in the other cases. Detailed descriptions of these process areas can be found in Section 3.3 of the Crude Methanol Study. Where there is case specific performance information, these features are presented in the relevant case sections.

3.2.1 Coal Milling, Grinding and Drying

The Shell process uses a dry feed system that is sensitive to the coal moisture content. For coal to flow smoothly through the lockhoppers that pressurize coal to the gasifier, the coal's surface moisture must be removed. The PRB coal used in this study contains 25.77% total moisture on an as-received basis. It is dried to 6% moisture for smooth flow through the dry feed system. The coal is simultaneously crushed and dried in the coal mill, then delivered to a surge hopper.

PRB coal is delivered to the site by 100-ton rail cars. It is unloaded into two receiving hoppers and fed to the vibratory feeder. It is then transferred through intermediate hoppers and silos to the coal crusher where it is reduced to 1-1/4" x 0 size.

3.2.2 Coal Preparation and Drying

The Shell process uses a dry feed system that is sensitive to coal moisture content. It was assumed that the coal must be dried to 6% moisture to allow for smooth flow through the dry feed system before feeding into the Shell entrained-flow gasifier. This moisture content is considered compatible with the storage, transport and feed injection requirements for the Shell entrained-flow gasifier.

The drying heat is provided by burning tail-gas from the Claus plant (for Cases 3a and 3c), flash and purge gas from the methanol synthesis process in an incinerator with air as the oxidant. In Nexant's simulation of the coal drying process, it was noted that after combining these fuel streams, there was still insufficient heat in Cases 3a and 3d to dry the PRB coal from 25.77% to 6% moisture. Supplemental natural gas therefore had to be fired to provide the remaining heat to dry the coal sufficiently.

The hot incinerator flue gas mixes with nitrogen from the ASU and a portion of the dryer exhaust gas in order to maintain a drying gas temperature of less than 500°F with oxygen content lower than 8%.

3.2.3 Air Separation Unit

All cases include an air separation unit (ASU) for generating oxygen. The ASU is a conventional, cryogenic, pumped liquid oxygen (LOX) unit that provides oxygen for the gasification process, as well as nitrogen for ancillary equipment. The ASU is designed to produce 95 mol% oxygen (O₂) for use in the gasifier and Claus plant. The air compressor is powered by an electric motor. Nitrogen is recovered and used as a diluent for coal drying, as described earlier in 3.2.2.

The battery limit conditions for the ASU products are summarized below:

Table 3-1
ASU Product Conditions

ASU Product	Pressure, psia	Temperature, °F
95% O ₂	23.2	55
Diluent N ₂	14.7	63
ASU Vent	16.4	64

3.2.4 Mercury Removal

Mercury removal is achieved by a packed bed of sulfur-impregnated activated carbon operating at 105°F. This packed-bed vessel is located upstream of the sulfur recovery unit with 20-second superficial gas residence time to achieve more than 90% removal of mercury in addition to removal of some portion of other volatile heavy metals such as arsenic. Mercury-removal systems using sulfur-impregnated activated carbon downstream of a coal gasifier have a reported bed life of 18 to 24 months, and usually replacement is required due to fouling of the bed rather than mercury saturation.

3.2.5 Slag and Ash Handling

Slag material drains from the gasifier into a water bath at the bottom of the gasifier vessel. The slag-water slurry is transferred to a slag crusher where the slag is crushed into pea size fragments. The slurry containing 5 to 10% solids is then transferred to a dewatering bin through a lockhopper. The separated water is clarified and reused as makeup to the water scrubber. The dried slag is stored for disposal.

3.2.6 Methanol Reactor and Synthesis Loop

The methanol synthesis route chosen for this study is the vapor-phase methanol process based on the breadth of operating experience with vapor-phase production units. The methanol reactor converts H₂ and CO to methanol in a packed-bed of catalyst. The primary side reactions produce ethanol, propanol and formaldehyde.

CO₂-lean syngas containing 2-3% CO₂ with a H₂/CO ratio of 2:1 from the Acid Gas Removal (AGR) process is compressed to the synthesis loop operating pressure of 755 psia in a syngas compressor. The compressed syngas is mixed with recycled gas, heated to 400°F, and routed to

the methanol reactor. The reactor is steam cooled to facilitate near isothermal operation at 475°F and 735 psia. In-line blowers, coolers and knock-out drums are used within the synthesis loop to maintain pressure and remove crude methanol.

In order to promote continuous production, the methanol reactor effluent is cooled to condense out the crude methanol that is removed in a phase separator. About 96% of the separated gas is compressed to reactor pressure and recycled along with fresh syngas to the methanol reactor. This recycling elevates the overall conversion of carbon, overcoming the low per-pass conversion of CO. A small purge gas stream (approximately 4%) is removed from the synthesis loop to limit the build-up of inert gas species. The purge gas is routed to the incinerator to be used as fuel for coal-drying.

3.2.7 Heat Recovery and Power Generation

Both the gasification and methanol synthesis processes generate a large amount of heat that can be recovered in the form of saturated steam, which is then used for either process requirements or power generation. The process steam is generated at three different pressure levels: high pressure (HP) steam at 2,415 psia, intermediate pressure (IP) steam at 360 psia and low pressure (LP) steam at 75 psia.

In the Crude Methanol Study, auxiliary power demand for the Case 2 CTM plant with CO₂ sequestration is met by two Rolls-Royce Trent 60 natural gas-fired turbines, each generating about 60 MWe of power. An additional 276 MWe is produced from the steam cycle, which uses a single reheat (2,400 psig/1,050°F/1,050°F) cycle. However, in Nexant's simulation of the power cycle, it was determined that the gas turbine flue gas outlet temperature was not high enough to superheat/reheat all the steam raised in the gasification and methanol synthesis section for power generation in the steam turbine.

Nexant's model of the power cycle uses three GE MS6001B gas turbines, each producing a nominal 45 MWe, and is more similar in operation to the cases from the DOE/NETL-2011/1477 Report (Cost and Performance Baseline for Fossil Energy Plants Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis).

A steam turbine was used to generate power from the steam generated in the Heat Recovery Steam Generator (HRSG) and process waste heat recovery (WHR) systems. HP steam at 2,415 psia and 900 °F and IP steam at 360 psia and 750 °F are used in the HP and IP stages of the steam turbine for power generation. There is no IP reheat in the HRSG as the gas turbine flue gas exhaust temperature is not high enough. LP exhaust steam from the last steam turbine stage is condensed by splitting 50/50 to a surface condenser and an air-cooled condenser to conserve cooling water. The condensers operate at 0.698 psia with a corresponding condensing temperature of 90 °F.

The condensates are collected and sent to a deaerator to remove dissolve gases and treated to provide BFW for the steam generators.

In order to achieve an overall CO₂ capture rate greater than 90%, the NGCC exhaust gas leaving the HRSG, which contains CO₂ from the combustion of natural gas, has to be cooled before

undergoing post-combustion capture. Fluor's Econamine FG+ process was used to capture 90% of the NGCC exhaust CO₂. The DOE/NETL 1397 report provides a more detailed description of this process.

3.2.8 Cooling Water Systems

Exhaust steam from the steam turbine is split 50/50 to a surface condenser cooled with cooling water and to an air-cooled condenser using ambient air and forced convection. The major impact of utilizing this parallel cooling method is a significant reduction in water requirement when compared to a wet cooling system.

The circulating water system is a closed-cycle cooling water system that supplies cooling water to the surface condenser to condense one-half of the main turbine exhaust steam. The system also supplies cooling water to the AGR plant as required, and to the auxiliary cooling system. The auxiliary cooling system is a closed-loop process that utilizes a higher quality water to remove heat from compressor intercoolers, oil coolers and other ancillary equipment and transfers that heat to the main circulating cooling water system in plate-and-frame heat exchangers. The heat transferred to the circulating water in the surface condenser and other applications is removed by a mechanical draft cooling tower.

The system consists of two 50 percent capacity vertical circulating water pumps, a mechanical draft evaporative cooling tower, and CS cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counter-flow mechanical draft cooling tower.

3.2.9 BFW/Condensate System

The function of the boiler feed water (BFW) system is to pump the various BFW streams from the deaerator storage tank in the HRSG to the respective steam drums. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow. The BFW pumps are supplied with instrumentation to monitor and alarm on low oil pressure, or high bearing temperature. BFW pump suction pressure and temperature are also monitored. In addition, the suction of each pump is equipped with a startup strainer.

3.2.10 Water Balance

Water required for the operation of the facility is obtained from a source such as a lake, river, or well. If the quality of the water is adequate, raw water is used directly as makeup to the cooling tower and the gasifier quench. To meet the rest of the plant's water needs, makeup must be treated first by filtration to create service-quality water. This quality of water serves as makeup to the plant's potable water, demineralized water, fire water, and service water systems. Higher quality boiler feedwater is treated by a typical reverse osmosis and electrodeionization package. Water rejected by the system is of an acceptable quality to be used as makeup to the cooling tower.

In addition to meeting the makeup water needs of the facility, water treatment systems must be capable of capturing and treating on-site waste streams. Wastewater created by the gasification process must pass through a number of pretreatment steps before being combined with other wastewater streams. Metals, ammonia, and suspended solids are removed from the stream through the use of a clarifier and a biological treatment unit. Once processed, the wastewater can be combined with the cooling tower blowdown as well as other plant waste streams in a final clarifier. Dechlorination and pH adjustment are performed as needed at this step of the process in order to meet all local discharge regulations. Solids separated out in this process are dried by means of a filter press and taken away for offsite disposal.

3.3 CASE 3a PROCESS DESCRIPTION

Case 3a is modeled upon process information provided in Case 2 of the Crude Methanol Report. The overall block flow diagram of the Case 3a reference plant is shown in Figure 3-1, with the accompanying stream flows shown in Table 3-2. Additional descriptions of the Case 3a plant's various processes are provided below. To better visualize the different unit operations in the CTM plant, simplified process flow diagrams (PFDs) of the various plant processes are depicted in Figures 3-2 through 3-9. Tables 3-3 through 3-10 provide the model-generated process data for the numbered streams referenced in the PFDs.

Coal Preparation and Drying

Coal receiving and handling is part of the Case 3a plant common areas and covered in Section 3.2.1. Coal is crushed in the coal mill and delivered to a surge hopper, which in turn delivers the coal to the coal pre-heater. The coal drying process, depicted in Figure 3-2 and described earlier in Section 3.2.2, reduces the PRB coal moisture content from 25.77 wt% to 6 wt%. The mass balances of the coal drying process are presented in Table 3-3.

Coal Feed System

The dried coal is drawn from the surge hoppers and fed through a pressurization lock hopper system to a dense phase pneumatic conveyor, which uses a stream of high pressure CO₂ (~800 psia) withdrawn from the CO₂ compressors to convey the coal to the gasifiers.

Air Separation Unit

The ASU process for the Case 3a reference plant is shown in Figure 3-3 with the gasification, quench and dry solid removal processes. The mass balance of this process is presented in Table 3-4.

The ASU's main air compressor is powered by an electric motor. Nitrogen is recovered and mainly used as a diluent in the coal drying process. A small, separate stream of nitrogen is compressed to be used as a stripping gas in the Rectisol[®] process.

Shell Gasifier

The gasification and quench processes are shown in Figure 3-3 and the mass balances are presented in Table 3-4. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables.

Syngas is generated from the gasification of PRB coal in a high-pressure, oxygen-blown Shell quench-type gasifier. The high-temperature entrained-bed gasifier uses a partial water quench and syngas cooler to cool the hot syngas stream and generate steam for the WGS reactors and power generation.

In the gasifier, the coal feedstock reacts with O_2 in a reducing environment to produce principally H_2 and CO with a small amount of CO_2 . High-temperature heat recovery in each gasifier train is accomplished in three steps, including the gasifier membrane wall, which maintains a protective ash layer over the membrane wall. The product gas from the gasifier is cooled using a syngas recycle quench to lower the temperature below the ash melting point. Syngas then goes through a raw gas cooler, which lowers the gas temperature and contributes to the production of HP steam for use in the steam cycle.

The solids are removed as both slag and ash. Liquid slag is solidified in a water bath and removed via a lock hopper system. Ash carried over with the syngas is removed in a ceramic candle filter. The collected ash is also removed via a lock hopper system. The syngas scrubber downstream of the gasifier removes any possibility of remaining PM passing the candle filter further downstream, by protecting against leakage from the filter seals or any undetected candle breakage that would allow large particulates into the scrubber.

The design size used in this study requires the use of eight operating trains with one spare train for a total of nine gasifiers. The facility contains one spare gasifier train to allow operation at 90% CF and to generally improve availability. The spare gasifier train feeds into the same gas clean-up trains as the other gasifier trains so that start-up/operation is transparent to the downstream processes.

Dry Solids Removal and Wet Scrubbing

The raw syngas exiting the ceramic particulate filter enters the scrubber for removal of chlorides and any remaining particulates. The quench scrubber washes the syngas in a countercurrent flow in two packed beds, which removes essentially all traces of entrained particles. The bottoms from the scrubber are sent to the slag removal and handling system for processing.

The dry solid removal processes are shown in Figure 3-3 with the ASU, gasification and quench processes. The mass balance for this process is shown in Table 3-3. The wet scrubber is shown in Figure 3-4, along with the WGS process and its mass balance is shown in Table 3-5.

Water Gas Shift

The WGS process is shown in Figure 3-3 and its material balance is shown Table 3-5. The stream numbers on the PFD correspond to the stream numbers in the mass balance table.

Coal-derived syngas from the wet scrubber enters the sour shift and cooling section. In order to achieve a 2:1 ratio of H₂ to CO in the final syngas to the methanol synthesis reactors, approximately 55 to 60 percent of the coal-derived syngas is shifted, with the remainder bypassing the shift reactors. Two shift reactors in series are used to achieve the desired H₂/CO composition in the syngas. Cooling is provided between the series of reactors to control the exothermic temperature rise. A gas-gas heat exchanger after the first WGS reactor is used to preheat the syngas prior to entering it, while also cooling the outlet gas before entering the second WGS reactor.

After the second-stage shift and subsequent cooling, the shifted syngas from the second-stage shift reactor outlet mixes with the bypass syngas and is further cooled before being sent to the downstream Rectisol[®] unit for sulfur and CO₂ removal.

Like the reference case in the Crude Methanol Study, the Case 3a syngas leaving the first WGS reactor exceeds 900°F. It should be noted that this condition potentially leads to side reactions that may deactivate the catalyst prematurely and change the product slate. Though not shown in this case, it may be necessary to inject additional steam, which will function as a heat carrier, to decrease the syngas temperature at the first WGS reactor outlet.

Low Temperature Gas Cooling and Mercury Removal

The low temperature gas cooling process, together with the mercury removal process, is shown in Figure 3-5 and its material balance is shown Table 3-6. The stream numbers on the PFD correspond to the stream numbers in the mass balance table.

Syngas is cooled in a number of steps to effectively recover heat to maximize efficiency. As the shifted syngas is cooled, IP and LP steam, process condensate and feed water are being heated.

Low pressure steam is used to strip NH₃ and other absorbed gases from the condensate in a sour water stripper. These stripped gases are sent to the Claus sulfur recovery unit to be treated with other sour gas streams. The stripped condensate mixes with process condensate separated from the syngas. The mixture is pumped and heated to about 390°F before being fed into the wet scrubbers.

Mercury removal is achieved via an activated carbon process described earlier in Section 3.2.4.

Acid Gas Removal

The AGR and CO₂ compression processes are shown in Figure 3-6, while the material balance is shown in Table 3-7. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table.

The CTM plant removes both H₂S and CO₂ within the same process via the Rectisol[®] unit. The Rectisol[®] AGR process was specified primarily because the methanol synthesis catalyst requires an H₂S level below 100 parts per billion by volume (ppbv) in order to maintain an adequate catalyst lifetime.

The Rectisol[®] process uses chilled methanol as a solvent. Because of the high vapor pressure of methanol, the process is operated at a temperature in the range of -30°F to -100°F. Although the reference Case 2 in the Crude Methanol Study provided the material balance for the Rectisol[®] process, this was only applicable for AGR operation at 515 psia. In Case 3c (AR gasifier-based CTM with Rectisol[®] AGR), the Rectisol[®] unit's operating pressure is higher since the AR gasification process is able to operate at higher pressures. There is no corresponding reference case that provides information on the Rectisol[®] process operating above 815 psia. With assistance from Linde, licensor of the Rectisol[®] process, HMBs for operating pressure at both 515 psia and 815 psia were estimated and the results were then used in Nexant's models to ensure a more accurate comparison between these two cases.

In the estimated HMB for Case 3c, the CO₂ product stream has half the CO concentration or 3,000 ppmv compared with the CO₂ product in the Case 2 of the Crude Methanol Study. This is still two orders of magnitude higher than the CO limit (35 ppmv) for saline reservoir CO₂ sequestration shown in Table 2-5, derived from the QGESS CO₂ Impurity Design Parameters document. However, from the same table, the CO concentration is within range stated in literature (10-5,000 ppmv) and is thus considered acceptable.

With input from Linde while the Rectisol[®] process' cost as reported in the NETL Crude Methanol Study is low lower its estimate is still considered to be within range, after accounting for the cost being reported in 2011 dollars. For Case 3a, the Rectisol[®] unit's cost estimate uses the Crude Methanol Study's Rectisol[®] cost.

CO₂ Compression

The CO₂ stream recovered from the Rectisol[®] unit is compressed to 2,215 psia in a multiple-stage, intercooled compressor to supercritical conditions. No drying is required since the CO₂ regenerated by the Rectisol[®] unit is free of water. Some CO₂ is withdrawn from this stream to be used for transporting the coal to the gasifier.

Sulfur Recovery Unit

The sulfur recovery unit (SRU) is shown in Figure 3-7, while the material balance is shown in Table 3-8. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table.

The purpose of the SRU is to treat the acid gas from the Rectisol[®] unit and sour gas streams from the sour water strippers to make an effluent gas acceptable for venting to the atmosphere or burning.

For this study, the SRU is a Claus bypass type SRU utilizing O₂ instead of air. The Claus plant produces molten sulfur by converting approximately one third of the H₂S in the feed to SO₂, then reacting the H₂S and SO₂ to sulfur and water. The combined Claus technology and tail gas recycle results in an overall sulfur recovery exceeding 99 percent. The liquid sulfur recovered goes to the sulfur pit, while the tail gas proceeds to the incinerator for coal drying.

Methanol Reactor and Synthesis Loop

The methanol synthesis process is shown in Figure 3-8, while the material balance is shown in Table 3-9. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table. The process follows the description stated earlier in Section 3.2.4.

A certain amount of makeup methanol will be required for continued operation of the Rectisol[®] AGR process. To account for this, the makeup methanol quantity to the Rectisol[®] unit was debited against the gross crude methanol production from the methanol synthesis unit. The net methanol production from the plant is therefore the methanol synthesis unit's gross production rate less the Rectisol[®] makeup quantity.

NGCC

Nexant's model of the power cycle uses three GE MS6001B gas turbines, each producing a nominal 45MW, and is more similar in operation to the cases from the DOE/NETL-2011/1477 Report (Cost and Performance Baseline for Fossil Energy Plants Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis).

A steam turbine was used to generate power from the steam generated in the HRSG and WHR systems. HP steam at 2,415 psia and 900 °F and IP steam at 360 psia and 750 °F are used in the HP and IP stages of the steam turbine for power generation. There is no IP reheat in the HRSG as the gas turbine flue gas hot enough. LP exhaust steam from the last steam turbine stage is condensed by splitting 50/50 to a surface condenser and an air-cooled condenser to conserve cooling water. The condensers operate at 0.698 psia with a corresponding condensing temperature of 90 °F.

The PFD for the NGCC section is shown in Figure 3-9, while the material balance is shown in Table 3-10. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table.

Power Cycle Flue Gas Post Combustion CO₂ Capture and Compression

As mentioned in Section 3.2.7, Fluor's Econamine FG+ process was used to capture CO₂ from the flue gas leaving the NGCC HRSG. Unlike the Rectisol[®] unit's regenerated CO₂, the CO₂ leaving the Econamine capture process still contains moisture. It is thus compressed separately to 2,215 psia by a multi-stage, intercooled centrifugal compressor that is equipped with a thermal swing adsorptive dryer, which dehydrates the CO₂ stream to a dew point of -40°F. The virtually moisture-free supercritical CO₂ stream is delivered to the plant B/L as sequestration ready.

Figure 3-1
Case 3a Reference Plant – Overall BFD

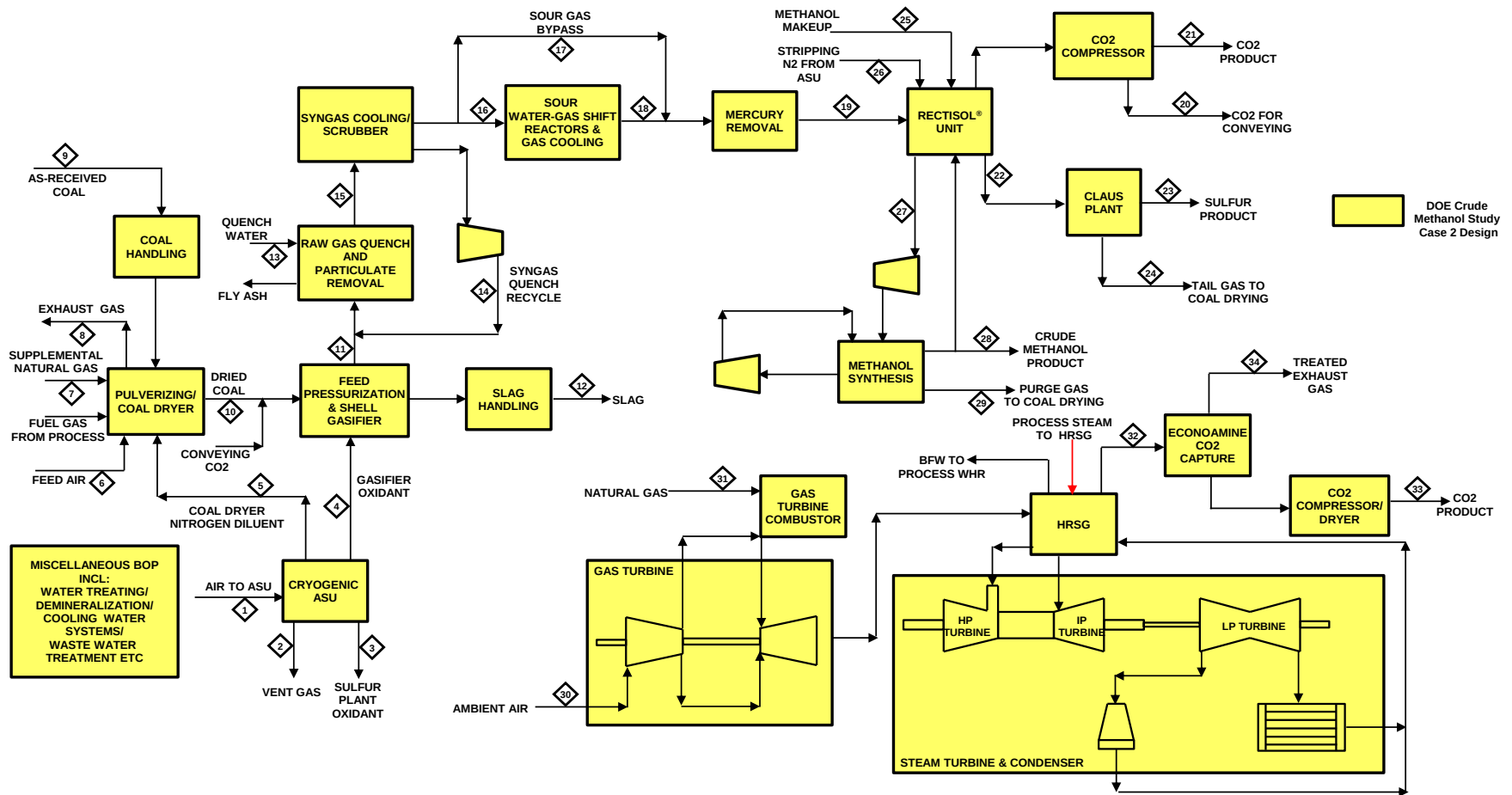


Table 3-2
Case 3a Reference Plant – Overall Stream Table

Description	Air to ASU	ASU Vent Gas	Sulfur Plant Oxidant	Gasifier Oxidant	Nitrogen Diluent to Dryer	Air Feed to Coal Dryer	Supplemental NG Feed	Coal Dryer Exhaust	As-Received Coal	Dried Coal	Raw Syngas	Slag	Quench Water	Cool Syngas Recycle	Syngas to Scrubber	Syngas to Shift	Bypass Syngas
Stream No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
V-L Mole Fraction																	
Ar	0.0092	0.0219	0.0300	0.0300	0.0023	0.0093	0	0.0087	0	0	0.0088	0	0	0.0047	0.0052	0.0047	0.0047
CH4	0	0	0	0	0	0	0.9310	0	0	0	0.0001	0	0	0.0000	0.0000	0.0000	0.0000
CO	0	0	0	0	0	0	0	0	0	0	0.6197	0	0	0.3319	0.3654	0.3319	0.3319
CO2	0.0003	0.0052	0	0	0	0	0	0	0	0	0.0639	0	0	0.0343	0.0377	0.0343	0.0343
COS	0	0	0	0	0	0	0	0	0	0	0.0005	0	0	0.0003	0.0003	0.0003	0.0003
H2	0	0	0	0	0	0	0	0	0	0	0.2313	0	0	0.1239	0.1364	0.1239	0.1239
H2O	0.0099	0.1535	0	0	0	0	0	0.1442	0	0	0.0616	0	1.0000	0.4973	0.4467	0.4973	0.4973
H2S	0	0	0	0	0	0	0	0	0	0	0.0030	0	0	0.0016	0.0018	0.0016	0.0016
N2	0.7732	0.6989	0.0200	0.0200	0.9921	0.7807	0.0160	0.8016	0	0	0.0111	0	0	0.0059	0.0065	0.0059	0.0059
NH3	0	0	0	0	0	0	0	0	0	0	0.0000	0	0	0	0	0	0
O2	0.2074	0.1206	0.9500	0.9500	0.0054	0.2095	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0.0000	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0.0320	0	0	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0	0.0070	0	0	0	0	0	0	0	0	0	0
n-Butane	0	0	0	0	0	0	0.0040	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0	0	1.0000	0	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	152690	9675	195	31278	110619	17089	130	154945	0	0	106160	0	61508	135448	303115	123016	75152
V-L Flowrate, lb/hr	4406168	264163	6273	1005820	3104006	495024	2252	4224734	0	0	2393030	0	1108084	2768779	6269875	2514637	1536237
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	1618190	1277850	0	136541	0	0	0	0	0
Temperature, F	59	64	72	292	70	59	80	156	59	156	2646		392	432	820	411	411
Pressure, psia	14.7	16.4	23.2	711.0	18.0	14.7	20.0	14.3	14.7	14.3	650.0	650.0	685.0	650.0	650.0	605.3	605.3

Table 3-2 (cont'd)
Case 3a Reference Plant – Overall Stream Table

Description	Shifted Syngas	Syngas to Rectisol®	Convey'g CO2	CO2 Product	Rectisol® Tail Gas to Claus	Sulfur Product	Tail Gas to Coal Drying	Methanol Makeup	Stripping N2 from ASU	Treated Syngas to MeOH	Crude Methanol Product	MeOH Synthesis Purge Gas	Air to Gas Turbine	Natural Gas Feed	NGCC Flue Gas Exhaust	CO2 Product	Treated NGCC Exhaust
Stream No.	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34
V-L Mole Fraction																	
Ar	0.0047	0.0069	0.0002	0.0002	0	0	0.0046	0	0.0023	0.0098	0.0001	0.1408	0.0092	0	0.0089	0	0.0096
CH4	0.0000	0.0000	0	0	0	0	0	0	0	0.0001	0	0.0008	0	0.9310	0	0	0
CO	0.0420	0.2220	0.0040	0.0040	0.0019	0	0.0085	0	0	0.3135	0	0.1798	0	0	0	0	0
CO2	0.3245	0.3131	0.9701	0.9701	0.6645	0	0.5659	0	0	0.0349	0.0114	0.2207	0.0005	0.0100	0.0351	1.0000	0.0038
COS	0.0000	0.0001	0	0	0	0	0.0012	0	0	0.0000	0	0	0	0	0	0	0
H2	0.4139	0.4439	0.0006	0.0006	0.0019	0	0.0037	0	0	0.6302	0	0.2655	0	0	0	0	0
H2O	0.2071	0.0027	0	0	0	0	0.3954	0	0	0	0.0119	0.0001	0.0099	0	0.0762	0	0.0332
H2S	0.0018	0.0025	0	0	0.2997	0	0.0028	0	0	0	0	0	0	0	0	0	0
N2	0.0059	0.0087	0.0246	0.0246	0.0142	0	0.0147	0	0.9921	0.0115	0.0001	0.1723	0.7730	0.0160	0.7471	0	0.8095
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0.0001	0.0001	0	0	0.0017	0	0.0054	0	0	0	0.2074	0	0.1328	0	0.1439
SO2	0	0	0	0	0	0	0.0014	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	1.0000	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0.0002	0	0.0002	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol	0	0	0.0004	0.0004	0	0	0	1.0000	0	0.0001	0.9751	0.0200	0	0	0	0	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0.0014	0	0	0	0	0	0
Propanol	0	0	0	0	0	0	0	0	0	0	0.0001	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0	0	0	0	0	0	0.032	0	0	0
Propane	0	0	0	0	0	0	0	0	0	0	0	0	0	0.007	0	0	0
n-Butane	0	0	0	0	0	0	0	0	0	0	0	0	0	0.004	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	123016	135645	5651	33904	1150	360	1402	22	923	95521	29303	5744	114330	3930	118374	3736	109248
V-L Flowrate, lb/hr	2514659	2923802	245924	1475527	47123	11529	46531	719	25907	1175263	938622	151588	3299430	68087	3367528	164443	3105993
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature, F	606	111	257	80	59	320	320	123	86	82	118	130	59	60	297	124	85
Pressure, psia	567.0	515.7	770.0	2215	28	20	20	40	75.0	485	40	717	14.5	474.7	14.9	2215	15.0

Figure 3-2
Case 3a Reference Plant – Coal Drying PFD

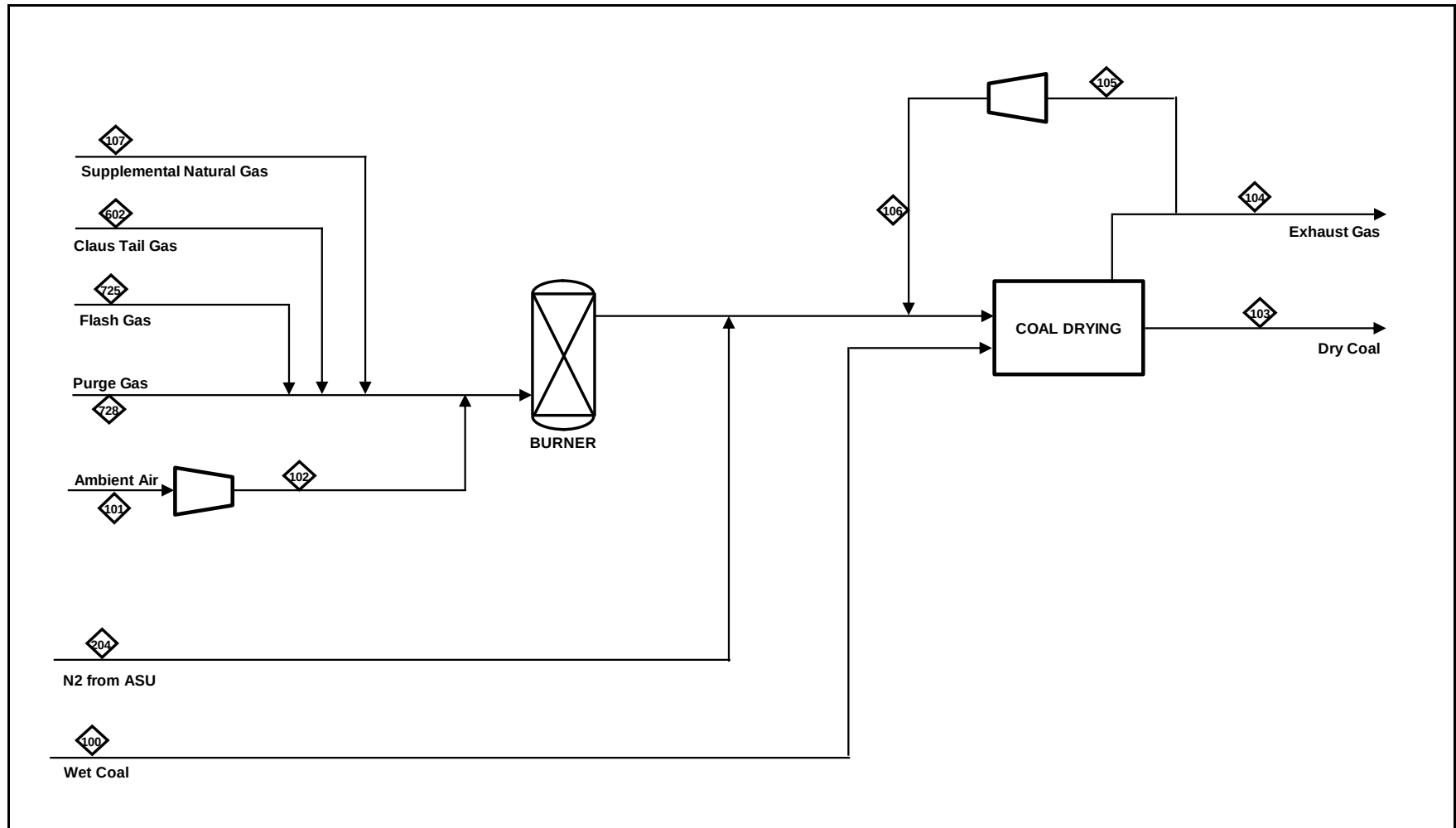


Table 3-3
Case 3a Reference Plant – Coal Drying Stream Table

STREAM	100	101	102	103	104	105	106	107	204	602	725	728
Description	Wet Coal	Ambient Air	Feed Air	Dry Coal	Exhaust Gas	LP Recycle Gas	HP Recycle Gas	Supplemental Nat Gas	Dilution Nitrogen	Claus Tail Gas	MeOH Flash Gas	MeOH Purge Gas
Mole Flow (Vapor/Liquid)/lbmol/hr												
AR	0	158	158	0	1350	727	727	0	254	6	122	809
CH4	0	0	0	0	0	0	0	121	0	0	1	4
CO	0	0	0	0	0	0	0	0	0	12	54	1033
CO2	0	6	6	0	5247	2825	2825	1	0	793	1425	1267
COS	0	0	0	0	0	0	0	0	0	2	0	0
H2	0	0	0	0	0	0	0	0	0	5	33	1525
H2O	0	169	169	0	22505	12118	12118	0	22	554	2	0
H2S	0	0	0	0	0	0	0	0	0	4	0	0
N2	0	13212	13212	0	124077	66811	66811	2	109745	21	108	990
NH3	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	3544	3544	0	1758	947	947	0	597	2	0	0
SO2	0	0	0	0	8	4	4	0	0	2	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0	0	0	398	115
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0	0.4	0.1
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	4	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	1	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	1	0	0	0	0
Total V/L Flow, lbmol/hr	0	17089	17089	0	154945	83432	83432	130	110619	1402	2143	5744
Total V/L Flow, lb/hr	0	493137	493137	0	4222847	2273841	2273841	2252	3104006	46531	85001	151587
Solids Mass Flow, lb/hr	1618190	0	0	1277850	0	0	0	0	0	0	0	0
Coal	1618190	0	0	1277850	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	1618190	493137	493137	1277850	4222847	2273841	2273841	2252	3104006	46531	85001	151587
Pressure, psia	14.7	14.7	17	14.3	14.3	14.3	16	20	18	20	40	717
Temperature, F		59	90		156	156	183	80	70	320	118	130

Figure 3-3
Case 3a Reference Plant – ASU/Coal Gasification PFD

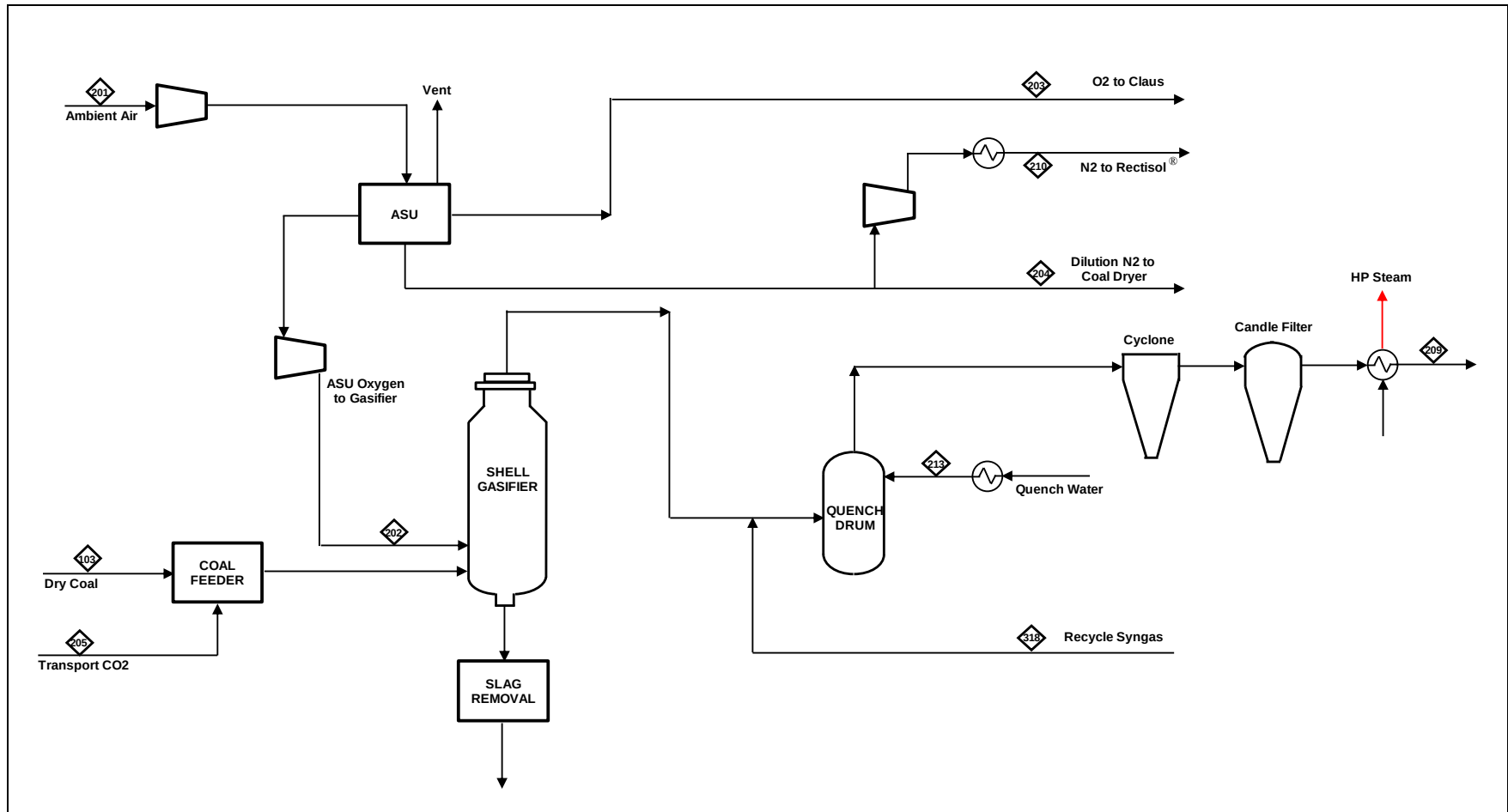


Table 3-4
Case 3a Reference Plant – ASU/Coal Gasification Stream Table

STREAM	201	202	203	204	205	103	318	213	214	209	210
Description	Ambient Air	Oxygen to Gasifier	Oxygen to Claus	N2 to Coal Dryer	Conveying CO2	Dry Coal	Recycle Syngas	Quench Water	Slag Out	Raw Cooled Syngas	Stripping N2 to Rectisol®
Mole Flow (Vapor/Liquid)/lbmol/hr											
AR	1412	938	6	254	1	0	642	0	0	1581	2
CH4	0	0	0	0	0	0	4	0	0	9	0
CO	0	0	0	0	23	0	44960	0	0	110745	0
CO2	50	0	0	0	5482	0	4646	0	0	11427	0
COS	0	0	0	0	0	0	35	0	0	86	0
H2	0	0	0	0	3	0	16783	0	0	41338	0
H2O	1507	0	0	22	0	0	67361	61508	0	135412	0
H2S	0	0	0	0	0	0	216	0	0	532	0
N2	118053	626	4	109745	139	0	802	0	0	1976	916
NH3	0	0	0	0	0	0	0	0	0	5	0
O2	31668	29714	185	597	1	0	0	0	0	0	5
SO2	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	1	0
HCL	0	0	0	0	0	0	0	0	0	3	0
Methanol -- CH3OH	0	0	0	0	2	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	152690	31278	195	110618.9	5651	0	135448	61508	0	303115	923
Total V/L Flow, lb/hr	4406168	1005820	6273	3104006	245924	0	2768779	1108084	0	6269875	25907
Solids Mass Flow, lb/hr											
Coal	0	0	0	0	0	1277850	0	0	136541	0	0
Ash	0	0	0	0	0	0	0	0	132490	0	0
Carbon	0	0	0	0	0	0	0	0	4051	0	0
Total Flow (Solids + V/L), lb/hr	4406168	1005820	6273	3104006	245924	1277850	2768779	1108084	136541	6269875	25907
Pressure, psia	14.7	711	23.2	18.0	770.0	14.3	650	685	650	605.3	75
Temperature, F	59	292	72	70	257		432	392		603	86

Figure 3-4
Case 3a Reference Plant – Wet Scrubber/Water Gas Shift PFD

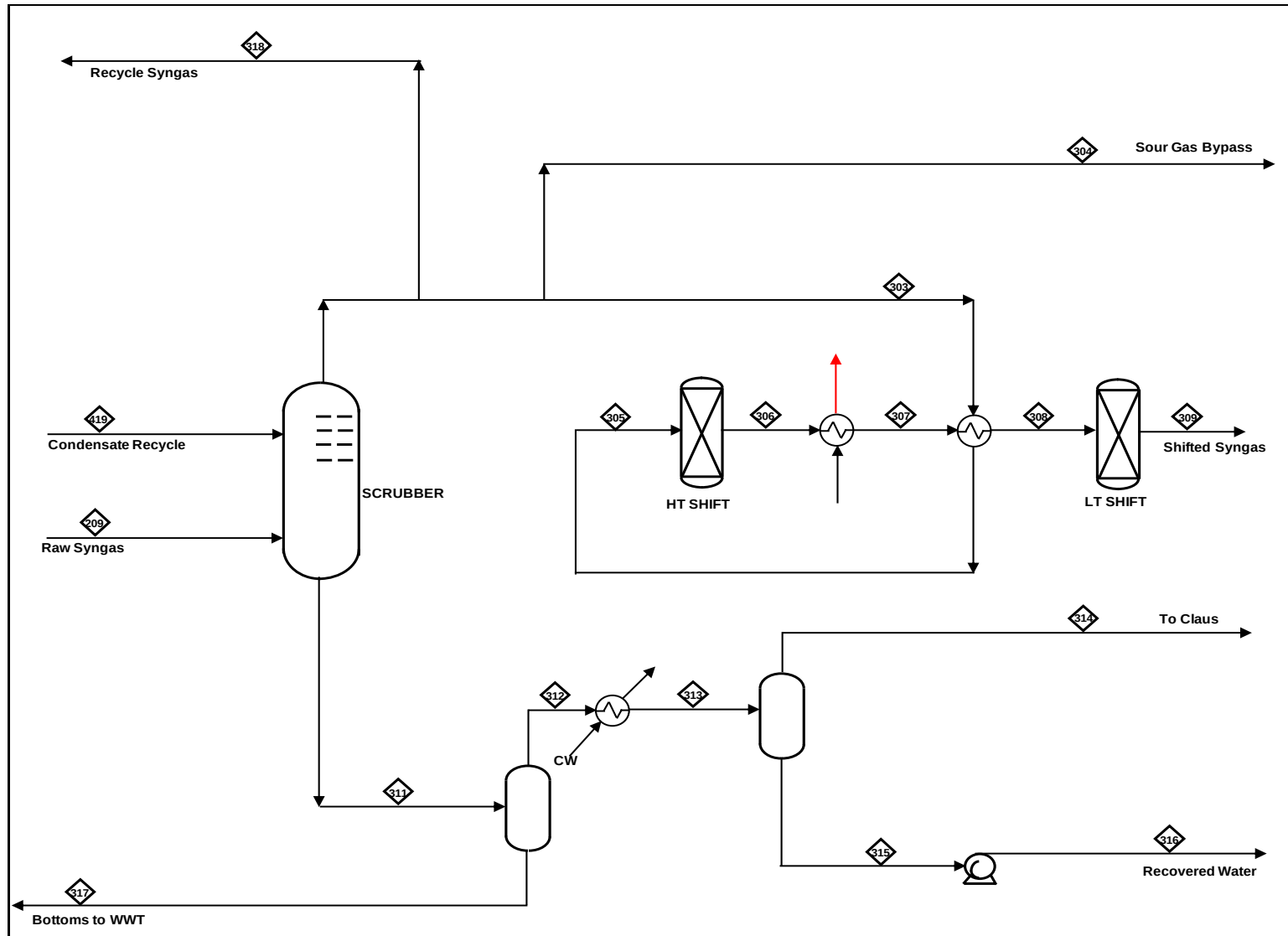


Table 3-5
Case 3a Reference Plant – Wet Scrubber/Water Gas Shift Stream Table

STREAM	209	419	303	304	305	306	307	308	309	311	312	313	314	315	317	318
Description	Raw Syngas	Conden sate Recycle	Syngas to Shift	Bypass Syngas	Hot Syngas	Hot 1st Shift Syngas	Warm 1st Shift Syngas	Cold 1st Shift Syngas	Hot 2nd Shift Syngas	Scrubber Bottoms	Scrubber Hot Vapor	Scrubber Cold Vapor	Scrubber Sour Gas	LP Recycle Water	Bottoms to WWT	Recycle Syngas
Mole Flow (Vapor/Liquid)/lbmol/hr																
AR	1581	0	583	356	583	583	583	583	583	1	1	1	1	0	0	642
CH4	9	0	3	2	3	3	3	3	3	0	0	0	0	0	0	4
CO	110745	1	40833	24945	40833	13206	13206	13206	5165	9	9	9	9	0	0	44960
CO2	11427	24	4219	2578	4219	31875	31875	31875	39918	8	8	8	8	0	0	4646
COS	86	0	32	19	32	3	3	3	1	0	0	0	0	0	0	35
H2	41338	3	15242	9312	15242	42869	42869	42869	50910	4	4	4	4	0	0	16783
H2O	135412	80532	61178	37375	61178	33521	33521	33521	25479	50031	7865	7865	1	7864	42166	67361
H2S	532	3	197	120	197	226	226	226	227	2	2	2	2	1	0	216
N2	1976	0	729	445	729	729	729	729	729	0	0	0	0	0	0	802
NH3	5	3	0	0	0	0	0	0	0	4	3	3	0	3	2	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCL	3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	303115	80566	123016	75152	123016	123016	123016	123016	123016	50059	7892	7892	24	7868	42168	135448
Total V/L Flow, lb/hr	6269875	1452070	2514637	1536237	2514637	2514654	2514654	2514654	2514659	902127.8	142467	142467	709.809	141758	759660	2768779
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	6269875	1452070	2514637	1536237	2514637	2514654	2514654	2514654	2514659	902128	142467	142467	710	141758	759660	2768779
Pressure, psia	605	762	605.3	605.3	600.3	593	584	575	567	600	51	46	46	46	51	605
Temperature, F	603	392	411	411	530	942	600	484	606	411	284	130	130	130	284	411

Figure 3-5
Case 3a Reference Plant – Low Temperature Gas Cooling PFD

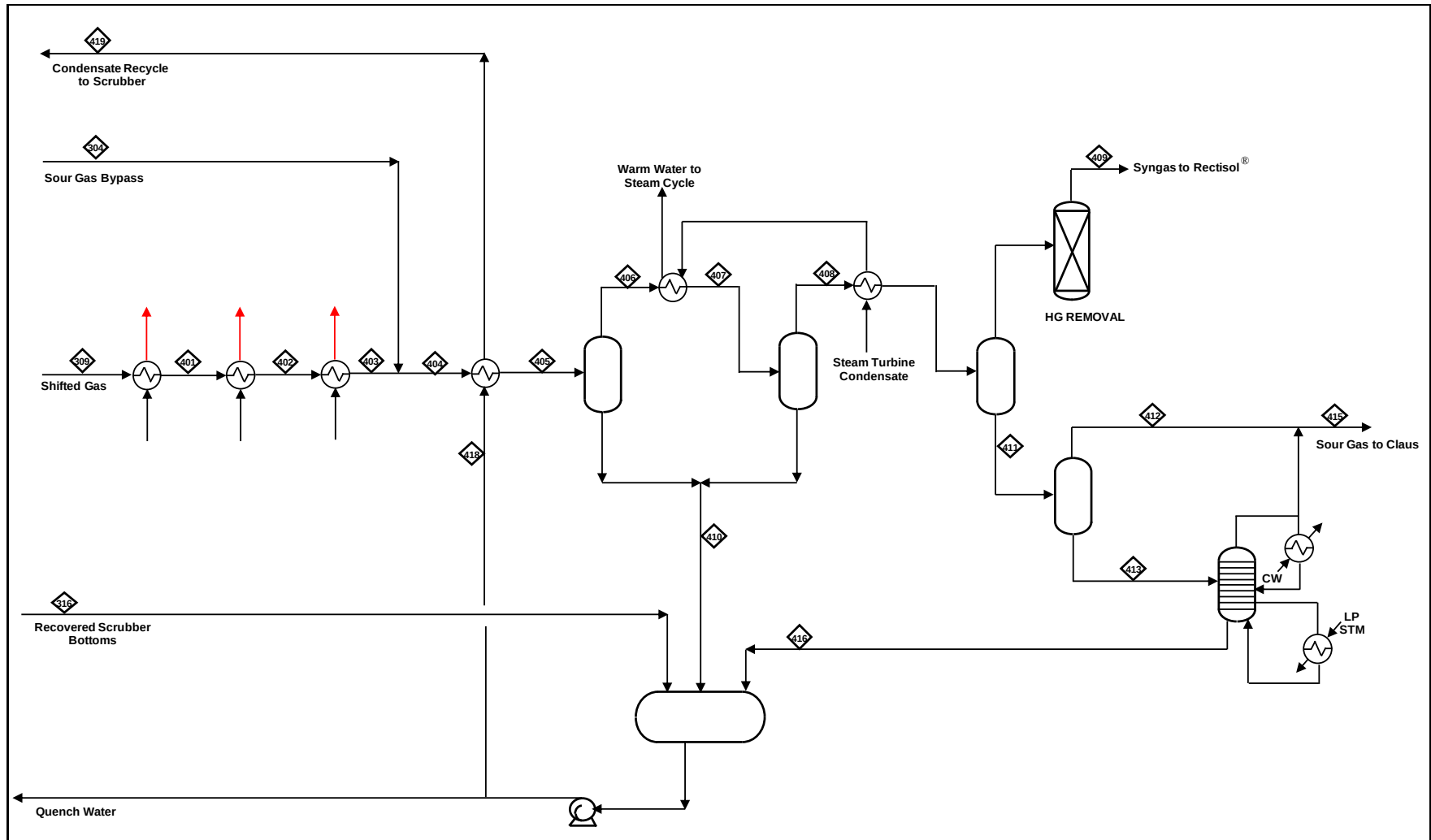


Table 3-6
Case 3a Reference Plant – Low Temperature Gas Cooling Stream Table

STREAM	309	403	304	404	405	406	407	408	409	415	416	316	419	213
Description	Hot 2nd Shift Syngas	Cooled Shift Syngas	Bypass Syngas	Mixed Syngas	Mixed Syngas Shift	1st KO Drum Vapor	Cool 1st KO Drum Vapor	2nd KO KO Drum Vapor	Syngas to Rectisol®	Sour Gas	Sour Stripper Bottoms	HP Recycle Water	Condensate Recycle	Quench Water
Mole Flow (Vapor/Liquid)/lbmol/hr														
AR	583	583	356	939	939	939	939	938	938	0	0	0	0	0
CH4	3	3	2	5	5	5	5	5	5	0	0	0	0	0
CO	5165	5165	24945	30111	30111	30110	30110	30109	30109	0	0	0	1	0
CO2	39918	39918	2578	42496	42496	42487	42487	42472	42468	4	0	0	24	0
COS	1	1	19	20	20	20	20	20	20	0	0	0	0	0
H2	50910	50910	9312	60222	60222	60221	60221	60219	60219	0	0	0	3	0
H2O	25479	25479	37375	62853	62853	49716	49716	23394	366	210	22819	7864	80532	61508
H2S	227	227	120	347	347	347	347	345	345	0	0	1	3	0
N2	729	729	445	1174	1174	1174	1174	1174	1174	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	3	3	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	123016	123016	75152	198168	198168	185020	185020	158678	135645	214	22819	7868	80566	61508
Total V/L Flow, lb/hr	2514659	2514659	1536237	4050895	4050895	3813804	3813804	3338853	2923802	3954.688	411096	141758	1452070	1108084
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	2514659	2514659	1536237	4050895	4050895	3813804	3813804	3338853	2923802	3955	411096	141758	1452070	1108084
Pressure, psia	567	552.69	605.3	552.7	549.7	544	540.69	540.69	515.69	61	74.16	46	762	685
Temperature, F	606	409	411	407	356	350	306	306	111	277	307	550	392	392

Figure 3-6
Case 3a Reference Plant – Rectisol® AGR and CO₂ Compression PFD

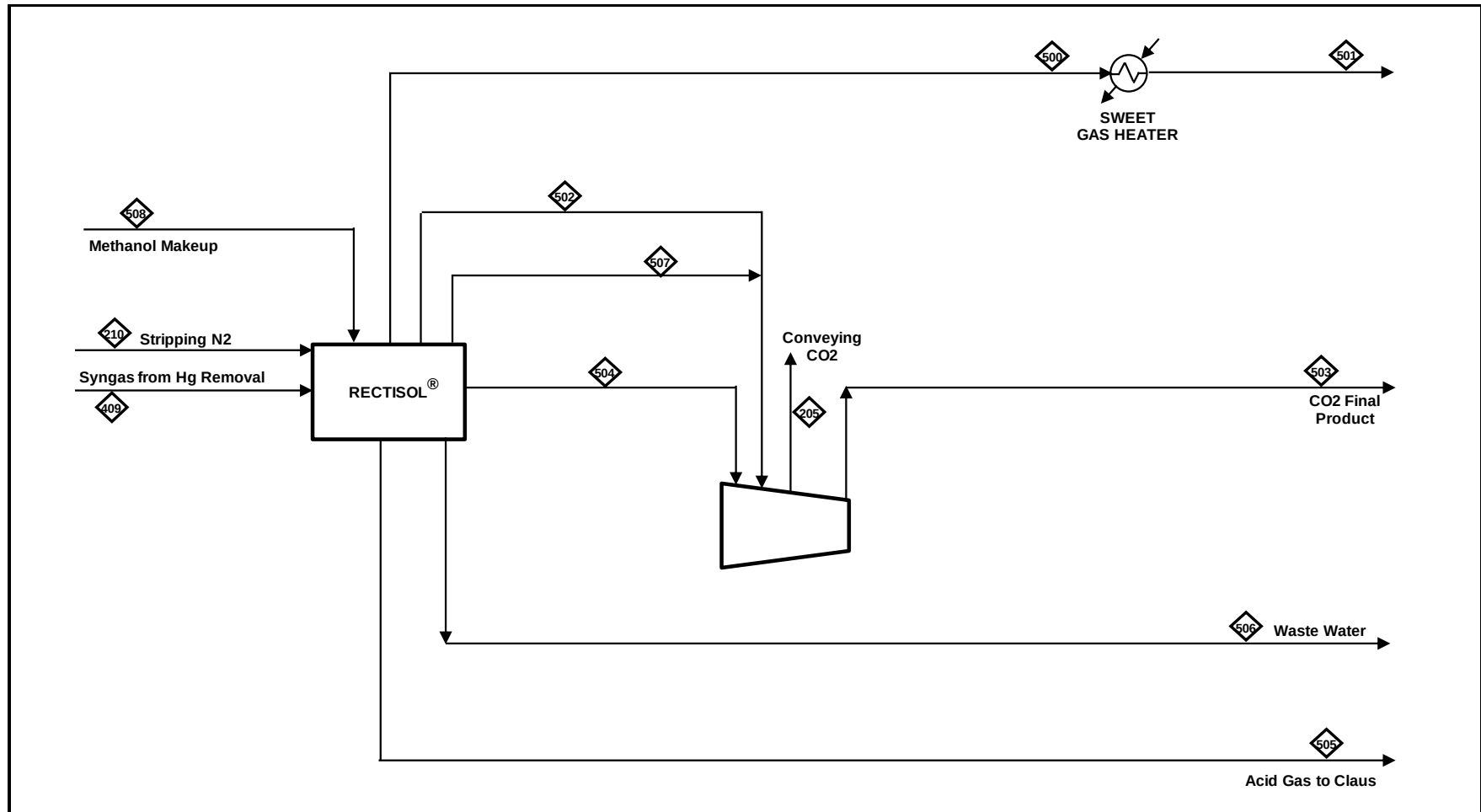


Table 3-7
Case 3a Reference Plant – Rectisol® AGR and CO₂ Compression Stream Table

STREAM	409	210	500	504	502	503	205	505	506	507	508
Description	Syngas to Rectisol®	Stripping N2 to Rectisol®	Treated Gas to MeOH	LP CO2 Product Stream	MP CO2 Product Stream	CO2 Final Product	Convey CO2	Acid Gas to Claus	Water to Waste	Fuel Gas to CO2 Product	Makeup Rectisol®
Mole Flow (Vapor/Liquid)/lbmol/hr											
AR	938	2	934	0	2	6	1	0	0	5	0
CH4	5	0	5	0	0	0	0	0	0	0	0
CO	30109	0	29948	2	76	137	23	2	0	81	0
CO2	42468	0	3333	30352	6418	32890	5482	764	0	1600	0
COS	20	0	0	0	0	0	0	20	0	0	0
H2	60219	0	60195	0	9	19	3	2	0	13	0
H2O	366	0	0	0	0	0	0	0	366	0	0
H2S	345	0	0	0	0	0	0	345	0	0	0
N2	1174	916	1099	970	2	835	139	16	0	2	0
NH3	0	0	0	0	0	0	0	0	0	0	0
O2	0	5	0	0	0	4	1	0	0	5	0
SO2	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	7	10	2	13	2	0	0	3	22
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	135645	923	95521	31334	6511	33904	5651	1150	366	1709	22
Total V/L Flow, lb/hr	2923802	25907	1175263	1363355	284865	1475527	245923.8	47123	6591	73231	719
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	2923802	25907	1175263	1363355	284865	1475527	245924	47123	6591	73231	719
Pressure, psia	516	75	485	19	42	2215	770	28	36	64	40
Temperature, F	111	86	82	81	81	162	257	59	261	81	123

Figure 3-7
Case 3a Reference Plant – Claus Plant PFD

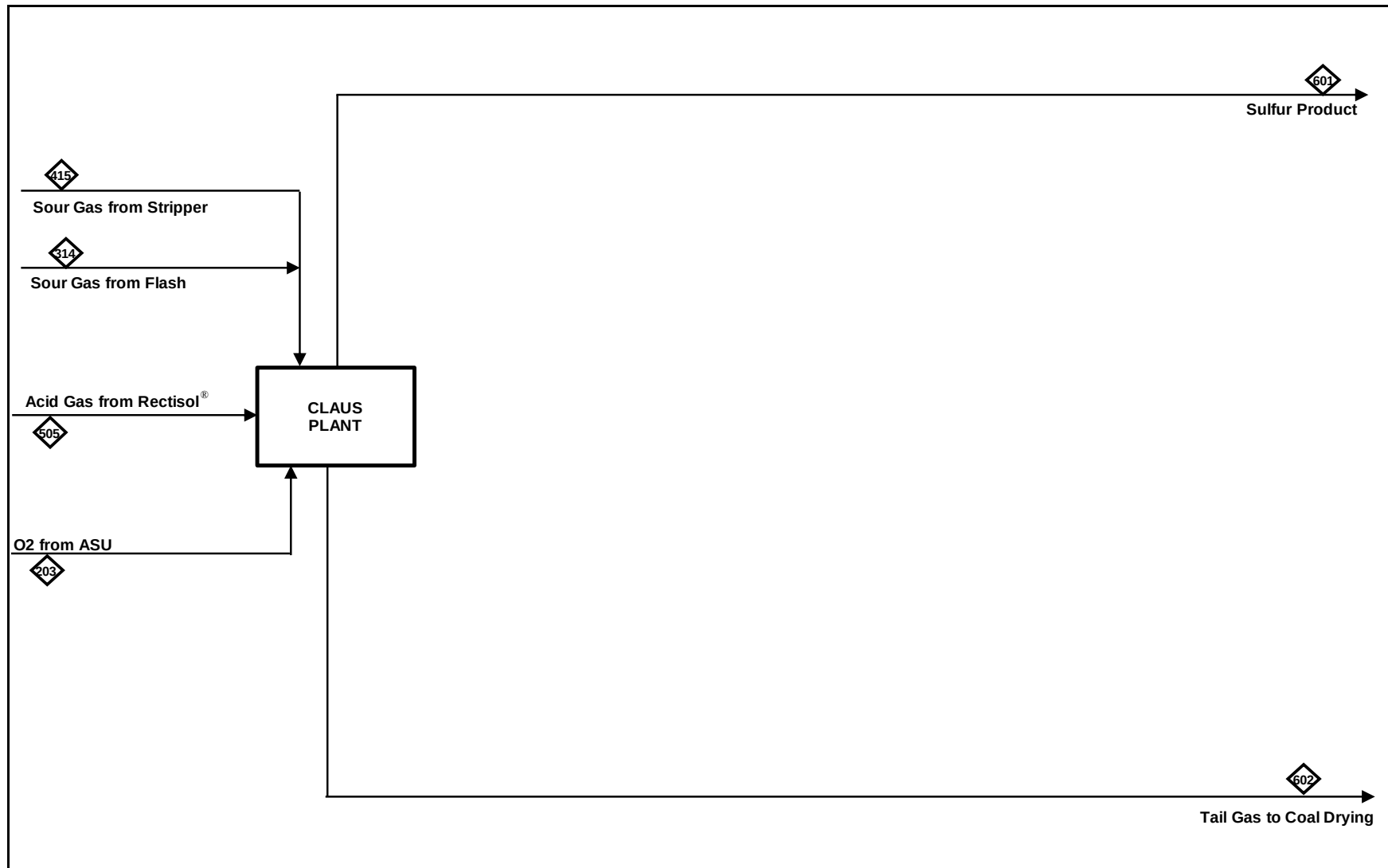


Table 3-8
Case 3a Reference Plant – Claus Plant Stream Table

STREAM	505	415	314	203	601	602
Description	Acid Gas from Rectisol®	Sour Gas from Stripper	Sour Gas from Flash	Claus O2 from ASU	Sulfur Product	Tail Gas to Coal Drying
Mole Flow (Vapor/Liquid)/lbmol/hr						
AR	0	0	1	6	0	6
CH4	0	0	0	0	0	0
CO	2	0	9	0	0	12
CO2	764	4	8	0	0	793
COS	20	0	0	0	0	2
H2	2	0	4	0	0	5
H2O	0	210	1	0	0	554
H2S	345	0	2	0	0	4
N2	16	0	0	4	0	21
NH3	0	0	0	0	0	0
O2	0	0	0	185	0	2
SO2	0	0	0	0	0	2
S	0	0	0	0	360	0
HCN	0	0	0	0	0	0
HCL	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	1150	214	24	195	360	1402
Total V/L Flow, lb/hr	47123	3955	710	6273	11529	46531
Solids Mass Flow, lb/hr	0	0	0	0	0	0
Coal	0	0	0	0	0	0
Ash	0	0	0	0	0	0
Carbon	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	47123	3955	710	6273	11529	46531
Pressure, psia	28	61	46	23	20	20
Temperature, F	59	277	130	72	320	320

Figure 3-8
Case 3a Reference Plant – Methanol Synthesis Plant PFD

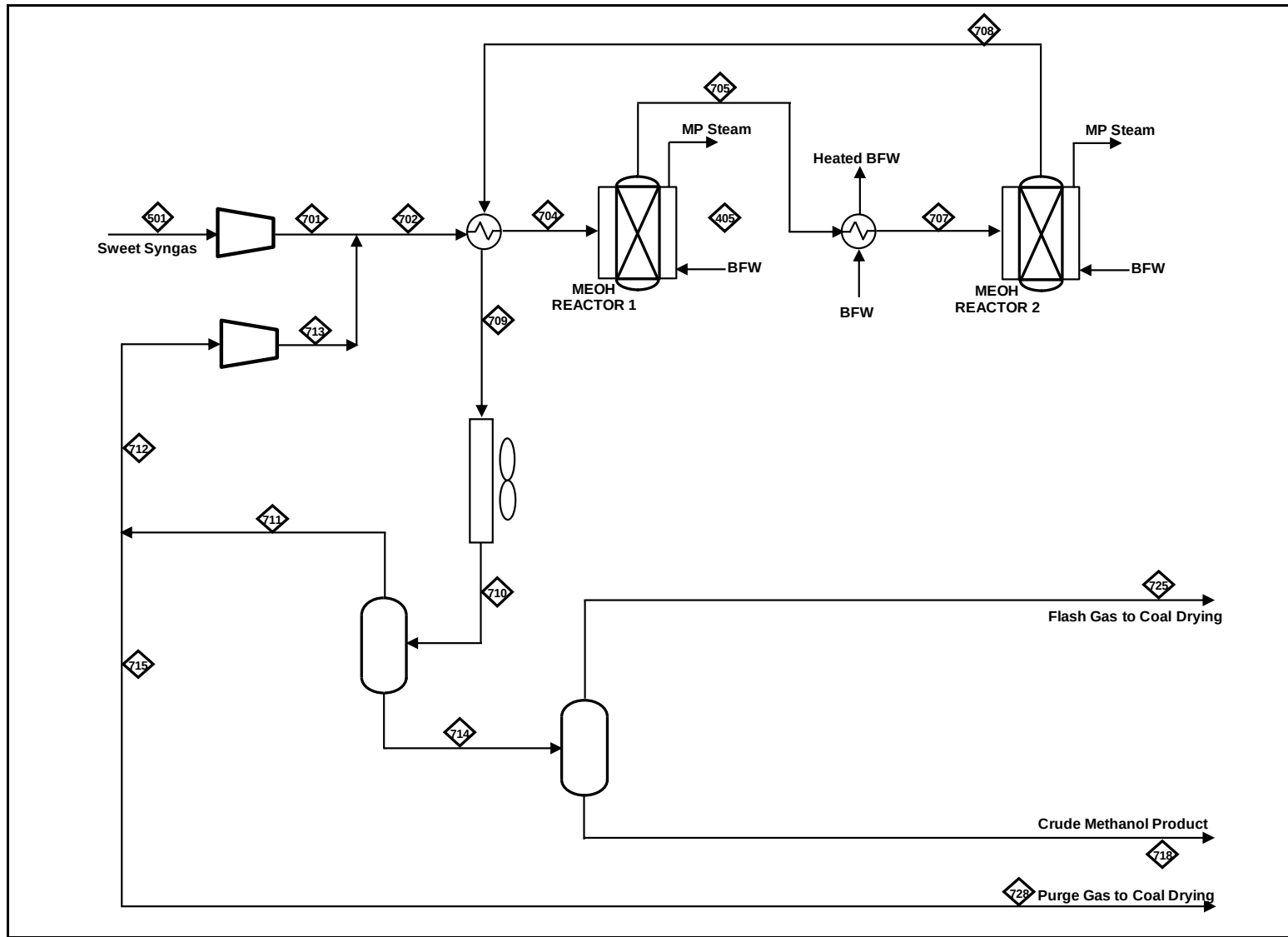


Table 3-9
Case 3a Reference Plant – Methanol Synthesis Plant Stream Table

STREAM Description	501 Reheated Syngas	701 HP Sweet Syngas	702 Mixed Syngas	704 MEOH Reactor 1 Feed	705 MEOH Reactor 1 Outlet	707 MEOH Reactor 2 Feed	708 MEOH Reactor 2 Outlet	709 Cooled Raw MeOH	710 AC Cool Raw MeOH	711 Flash Gas Ovhd	712 Recycle Syngas	713 HP Recycle Syngas	714 KO Drum Bottoms	718 Crude MeOH Product	725 MeOH Flash Gas	728 MeOH Purge Gas
Mole Flow (Vapor/Liquid)/lbmol/hr																
AR	934	934	20344	20344	20344	20344	20344	20344	20344	20219	19410	19410	125	3	122	809
CH4	5	5	109	109	109	109	109	109	109	108	104	104	1	0	1	4
CO	29948	29948	54739	54739	35143	35143	25878	25878	25878	25824	24791	24791	54	0	54	1033
CO2	3333	3333	33753	33753	33214	33214	33448	33448	33448	31687	30420	30420	1761	335	1425	1267
COS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	60195	60195	96796	96796	55987	55987	38159	38159	38159	38126	36601	36601	33	0	33	1525
H2O	0	0	11	11	595	595	362	362	362	12	11	11	350	348	2	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	1099	1099	24851	24851	24851	24851	24851	24851	24851	24742	23752	23752	109	2	108	990
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	7	7	2766	2766	22814	22814	31844	31844	31844	2874	2759	2759	28970	28572	398	115
Ethanol -- C2H5OH	0	0	3	3	43	43	44	44	44	3	3	3	41	41	0	0
Propanol -- C3H8O	0	0	0	0	2	2	2	2	2	0	0	0	2	2	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	95520.94	95521	233371	233371	193101	193101	175040	175040	175040	143595	137850	137850	31445	29303	2143	5744
Total V/L Flow, lb/hr	1175263	1175263	4813379	4813379	4813343	4813343	4813327	4813327	4813327	3789704	3638116	3638116	1023623	938622	85001.3	151588
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	1175263	1175263	4813379	4813379	4813343	4813343	4813327	4813327	4813327	3789704	3638116	3638116	1023623	938622	85001	151588
Pressure, psia	483	755	755	747.0	737.0	732.0	727	722	720	717	717	755	717	40	40	717
Temperature, F	125	234	175	400	475	400	430	239	130	130	130	141	130	118	118	130

Figure 3-9
Case 3a Reference Plant -- NGCC PFD

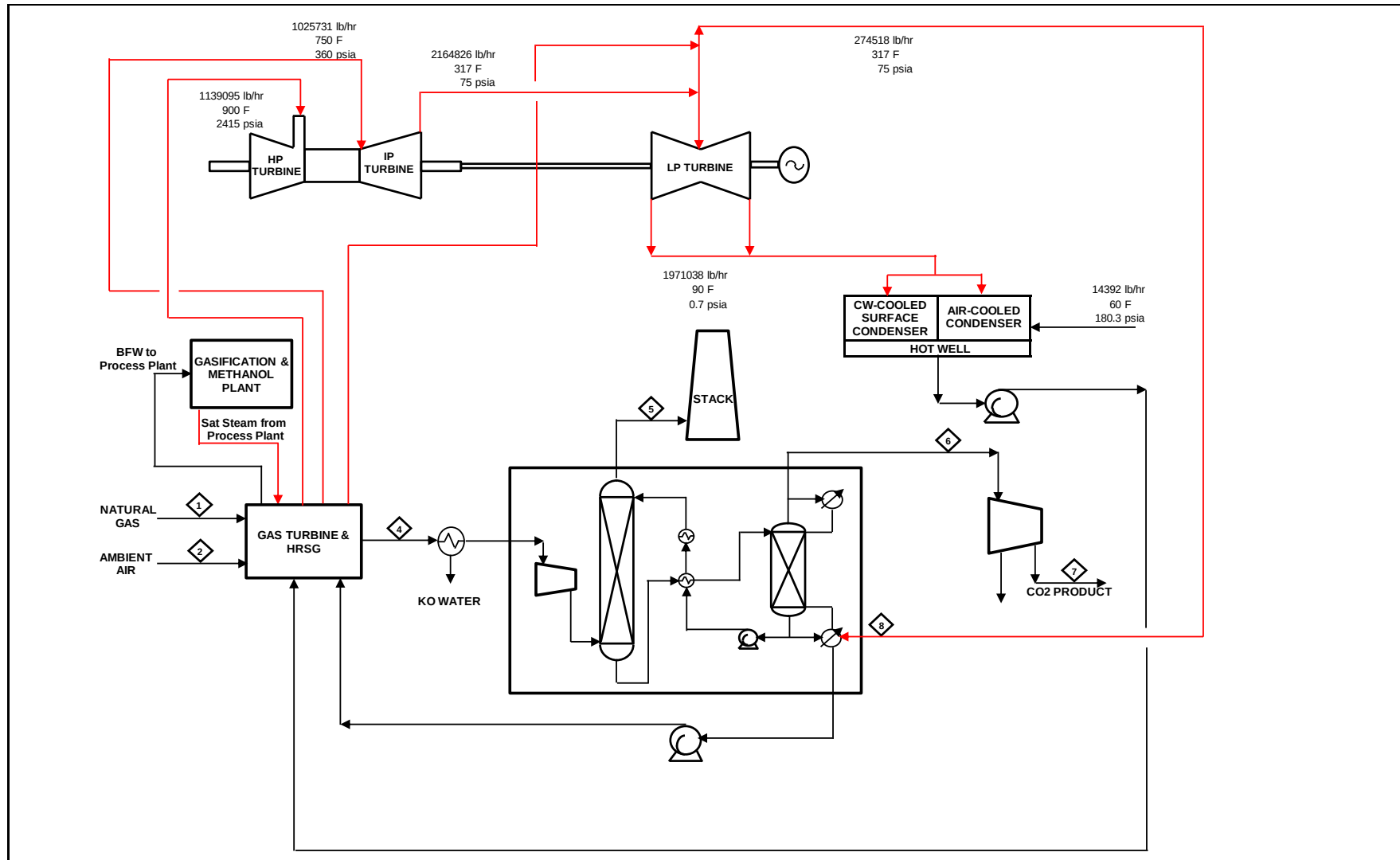


Table 3-10
Case 3a Reference Plant – NGCC Stream Table

STREAM	1	2	4	5	6	7	8
Description	Ambient Air	Natural Gas	Flue Gas to MEA	Treated Gas	CO2 Out	CO2 Product	LP Steam
Mole Flow (Vapor/Liquid)/lbmol/hr							
AR	1053	0	1053	1053	0	0	0
CH4	0	3659	0	0	0	0	0
CO	0	0	0	0	0	0	0
CO2	57	39	4152	415	3736	3736	0
COS	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0
H2O	1132	0	9016	3626	45	0	15261
H2S	0	0	0	0	0	0	0
N2	88373	63	88436	88436	0	0	0
NH3	0	0	0	0	0	0	0
O2	23715	0	15717	15717	0	0	0
SO2	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0
Ethane -- C2H6	0	126	0	0	0	0	0
Propane -- C3H8	0	28	0	0	0	0	0
n-Butane -- C4H10	0	16	0	0	0	0	0
Total V/L Flow, lbmol/hr	114330	3930	118374	109248	3782	3736	15261
Total V/L Flow, lb/hr	3299430	68087.15	3367528	3105993	165255.9	164443.1	274940.1
Total Flow (Solids + V/L), lb/hr	3299430	68087	3367528	3105993	165256	164443	274940
Pressure, psia	14.5	474.7	15	15	24	2215	260
Temperature, F	59	60	297	85	69	124	35

3.4 SPARING PHILOSOPHY

The design size requires eight gasifier trains with one spare train for a total of nine gasifiers (9 x 12.5%). No further sparing information was provided in the Crude Methanol Study for the rest of the CTM plant's systems.

3.5 PERFORMANCE RESULTS

The Nexant-modeled Case 3a reference plant with CO₂ capture consumes 19,418 tpd of PRB coal and 37.0 million standard cubic feet per day (MMSCFD) of natural gas at the Midwestern site and produces 937,909 lb/hr (10,210 mtpd) of crude methanol on a net basis. Overall performance for the Case 3a reference plant is summarized in Table 3-11, which includes auxiliary power requirements.

Table 3-11
Case 3a Reference Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case 3a
Gas Turbine Power	127,087
Steam Turbine Power	264,700
TOTAL POWER, kWe	391,787
CTM Plant Auxiliary Load Summary, kWe	
Coal Handling	9,090
Slag Handling	2,377
Incinerator Air Blower	4,875
Incinerator Recycle Blower	1,090
Air Separation Unit	175,570
Quench Water Pump	1,661
Syngas Recycle Compressor	6,418
Scrubber Pumps	1,068
Flash Bottoms Pump	570
Rectisol® AGR Auxiliaries	42,263
Claus Plant Auxiliaries	249
CO ₂ Compressor Auxiliaries	65,213
Methanol Synthesis Syngas Feed Compressor	22,257
Recycle Gas Compressor	3,289
Air Cooler Fans	1,786
Water Treatment	3,046
Miscellaneous BOP	5,000
Circulating Water Pumps	8,195
Ground Water Pumps	873
Cooling Tower Fans	5,343
SUBTOTAL CTM PLANT AUXILIARIES, kWe	360,233
NGCC Plant Auxiliary Load Summary, kWe	
Gas Turbine Auxiliaries	782
Steam Turbine Auxiliaries	100
Transformer Losses	2,149
Miscellaneous BOP	500
Air Cooled Condenser Fans	4,135
Condensate Pumps	748
Boiler Feed Water Pumps	13,967
Amine System Auxiliaries	3,893
NGCC CO ₂ Compression	6,033
SCR	16
SUBTOTAL POWER PLANT AUXILIARIES, kWe	32,324
TOTAL AUXILIARIES, kWe	392,55
NET PLANT POWER, kWe	-771

Table 3-12 shows the carbon balance for the Case 3a reference plant. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal and natural gas feedstock. Carbon in the air is not part of the carbon capture equation, but is not neglected in the balance since the model accounts for the air components throughout. Carbon leaves the CTM process plant as unburned carbon in the slag, CO₂ in the dryer exhaust gas, ASU vent gas, crude methanol product, and the Rectisol® AGR CO₂ capture product. Additionally, carbon also leaves the power cycle as CO₂ in the stack and Econamine CO₂ capture product. Carbon in the crude methanol is considered as product, not emissions. The carbon capture efficiency is defined as the amount of carbon in the product streams, which include the crude methanol and the dried and compressed CO₂ products from the Rectisol® AGR and Econamine CO₂ capture processes, relative to the amount of carbon in the coal and natural gas feedstock less carbon contained in the slag. For Case 3a, the carbon capture efficiency is 92.1%.

Table 3-12
Case 3a Reference Plant – Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	810,195	
ASU Air	602	
Coal Dryer Air	67	
Coal Dryer Supplemental Natural Gas	1,627	
Power Cycle Natural Gas	49,186	
Power Cycle Combustion Air	450	
ASU Vent		602
Coal Dryer Exhaust Gas		63,020
Carbon in Slag		4,051
Sulfur Product		0
Crude Methanol Product		347,987
Rectisol® AGR CO ₂ Product		396,831
NGCC CO ₂ Product		44,672
NGCC Exhaust Gas		4,964
Total	862,127	862,127

Table 3-13 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur emitted in the coal dryer exhaust gas, and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible. The net sulfur emissions include only the sulfur emitted in the coal dryer exhaust gas. Based on this, the net sulfur emissions for the plant are 0.035 lb of SO₂/MMBtu, which meet the sulfur emission limit in the BACT environmental design basis in Table 2-7. However, this sulfur emission exceeds the environmental target used in the companion IGCC studies performed on the advanced technologies. Actual permitting for the plant will also include the methanol process in addition to the gasification and power generation, which generally targets lower sulfur emissions. If the permit for this plant requires lower sulfur emissions, additional sulfur removal would be required for this plant.

Table 3-13
Case 3a Reference Plant – Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	11,772	
Sulfur Product		11,530
Coal Dryer Exhaust Gas		241
CO ₂ Product		0
Convergence Tolerance		1
Total	11,772	11,772

Table 3-14 shows the overall water balance for the Case 3a plant. Raw water withdrawal is the amount of raw water consumed by the plant. The raw water is obtained from groundwater (50%) and from municipal sources (50%). Some water is discharged by the processes as effluent suitable for internal consumption. This effluent is internally recycled and consumed by the slag handling process, as make up to the syngas scrubber, and as makeup to the cooling tower. Some water is discharged from the process to a permitted outfall. Waste water is discharged by the slag handling process, the Rectisol® unit, and as cooling tower blowdown.

Table 3-14
Case 3a Plant – Overall Water Balance

Water Use, gpm	Raw Water Withdrawal	Process Effluent Production for Internal Consumption	Internal Consumption	Process Water Discharge
Slag Handling	0	0	353	(353)
Quench Cooler	2,214	0	0	0
Syngas Scrubber Consumption	374	0	2,525	0
Syngas Scrubber Effluent	0	(1,801)	0	0
Syngas Cooling Knockout	0	(1,420)	0	0
Sour Water Stripper Effluent	0	(821)	0	0
Rectisol Unit	0	0	0	(13)
Steam Cycle Makeup	29	0	0	0
Steam Cycle Blowdown	0	(29)	0	0
Flue Gas Cooling Knockout	0	(192)	0	0
CO ₂ Compression Knockout	0	(2)	0	0
Cooling Tower Makeup	6,969	0	1,387	0
Cooling Tower Blowdown	0	0	0	(1,895)
Total	9,586	(4,265)	4,265	(2,261)

Positive values represent consumption while negative values represent production

3.6 EQUIPMENT LIST

No equipment list was provided for the reference case in the Crude Methanol Study. However, the reference case systems that warrant consideration in this study (gasification, syngas cleanup and CO₂ capture systems) are similar to the ones listed in Case S1B in the *Cost and Performance*

Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity: IGCC Cases, DOE/NETL-2010/1399 report (NETL Report 1399). The reader should hence refer to the Case S1B equipment list in NETL Report 1399.

3.7 CAPITAL COST

Table 3-15 shows the cost breakdown of the Case 3a reference plant, consistent with the code of accounts format as expressed in the Crude Methanol Study. The accounts/subaccounts of interest for this study are:

- 2.3 Dry Coal Injection System,
- 4 Gasifier & Accessories,
- 5A Gas Cleanup & Piping and
- 5B CO₂ Removal and Compression

These are shown with more detail to include the various subaccounts and provide more clarity to the major cost differences among the cases.

Table 3-16 shows the calculation and addition of owner's costs to determine the TOC, which is used to calculate the product methanol RSP.

The estimated TOC of the Case 3a reference plant in 2011 dollars is \$5,892MM.

Table 3-15
Case 3a CTM Plant – Total Plant Cost Summary

Case 3a: Shell with Rectisol® AGR CTM Plant Low Rank Western Coal Baseline Study			PRB		Coal Feed, lb/hr Coal HHV, Btu/lb		1,618,190 8,564		Plant Size		10210 metric tons per day	
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost \$	Eng'g CM H.O & Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect				Process	Project	\$	
1	COAL & SORBENT HANDLING	\$36,523	\$6,549	\$28,311	\$0	\$0	\$71,383	\$6,479	\$0	\$15,572	\$93,434	
2	COAL & SORBENT PREP & FEED											
2.3	Dry Coal Injection System	\$203,470	\$0	\$137,594	\$0	\$0	\$341,064	\$29,375	\$0	\$74,088	\$444,527	
2.x	Other Coal & Sorbent Prep & Feed Systems	\$151,813	\$28,795	\$43,835	\$0	\$0	\$224,443	\$19,633	\$0	\$48,816	\$292,892	
	SUBTOTAL 2.	\$355,283	\$28,795	\$181,429	\$0	\$0	\$565,507	\$49,008	\$0	\$122,904	\$737,419	
3	FEEDWATER & MISC BOP SYSTEMS	\$29,758	\$8,475	\$18,183	\$0	\$0	\$56,416	\$5,299	\$0	\$14,019	\$75,734	
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries (Shell)	\$429,138	\$0	\$194,177	\$0	\$0	\$623,315	\$55,887	\$89,766	\$118,739	\$887,707	
4.3	ASU/Oxidant Compression	\$602,252	\$0	\$0	\$0	\$0	\$602,252	\$58,376	\$0	\$66,062	\$726,690	
4.4	LT Heat Recovery & FG Saturation	\$88,402	\$0	\$40,000	\$0	\$0	\$128,403	\$12,531	\$0	\$28,187	\$169,121	
4.6	Flare Stack System	\$0	\$1,462	\$666	\$0	\$0	\$2,128	\$204	\$0	\$466	\$2,798	
4.9	Gasification Foundations	\$0	\$31,373	\$18,710	\$0	\$0	\$50,083	\$4,585	\$0	\$13,667	\$68,335	
	SUBTOTAL 4.	\$1,119,792	\$32,835	\$253,553	\$0	\$0	\$1,406,181	\$131,583	\$89,766	\$227,121	\$1,854,651	
5A	GAS CLEANUP & PIPING											
5A.1	Rectisol® System	\$353,257	\$0	\$0	\$0	\$0	\$353,257	\$33,374	\$70,651	\$91,457	\$548,738	
5A.2	Elemental Sulfur Plant	\$10,010	\$1,951	\$12,826	\$0	\$0	\$24,787	\$2,407	\$0	\$5,439	\$32,633	
5A.3	Mercury Removal	\$7,601	\$0	\$5,745	\$0	\$0	\$13,346	\$1,289	\$667	\$3,060	\$18,362	
5A.4	Shift Reactors	\$13,343	\$0	\$5,333	\$0	\$0	\$18,677	\$1,790	\$0	\$4,094	\$24,561	
5A.5	Syngas Compressor	\$6,426	\$0	\$4,864	\$0	\$0	\$11,290	\$1,075	\$0	\$1,237	\$13,603	
5A.6	Blowback Gas Systems	\$6,751	\$1,136	\$640	\$0	\$0	\$8,527	\$809	\$0	\$1,867	\$11,203	
5A.7	Fuel Gas Piping	\$0	\$2,917	\$1,908	\$0	\$0	\$4,825	\$447	\$0	\$1,054	\$6,326	
5A.9	HGCU Foundations	\$0	\$2,644	\$1,782	\$0	\$0	\$4,426	\$407	\$0	\$1,450	\$6,283	
	SUBTOTAL 5.	\$397,388	\$8,648	\$33,099	\$0	\$0	\$439,134	\$41,599	\$71,318	\$109,658	\$661,709	
5B	CO2 REMOVAL & COMPRESSION											
5B.1	NGCC CO2 Removal System	\$22,922	\$0	\$6,907	\$0	\$0	\$29,830	\$2,489	\$5,966	\$7,658	\$45,942	
5B.2	CO2 Compression & Drying (Rectisol)	\$67,218	\$0	\$25,340	\$0	\$0	\$92,558	\$8,868	\$0	\$20,285	\$121,711	
5B.3	CO2 Compression & Drying (NGCC)	\$10,028	\$0	\$3,839	\$0	\$0	\$13,867	\$1,160	\$0	\$3,005	\$18,032	
	SUBTOTAL 5B.	\$100,168	\$0	\$36,086	\$0	\$0	\$136,254	\$12,517	\$5,966	\$30,948	\$185,685	
5C	METHANOL PRODUCTION	\$132,878	\$55,904	\$111,809	\$0	\$0	\$300,591	\$30,060	\$0	\$66,130	\$396,781	
6	COMBUSTION TURBINE/ACCESSORIES	\$47,397	\$394	\$3,814	\$0	\$0	\$51,605	\$11,171	\$11,909	\$14,852	\$89,537	
7	HRSG, DUCTING & STACK	\$29,304	\$819	\$6,961	\$0	\$0	\$37,083	\$3,079	\$0	\$4,038	\$44,201	
8	STEAM TURBINE GENERATOR	\$101,175	\$1,098	\$25,478	\$0	\$0	\$127,752	\$11,474	\$0	\$27,159	\$166,385	
9	COOLING WATER SYSTEM	\$18,209	\$19,919	\$14,802	\$0	\$0	\$52,930	\$4,897	\$0	\$11,755	\$69,582	
10	ASH/SPENT SORBENT HANDLING SYS	\$82,188	\$3,649	\$77,402	\$0	\$0	\$163,239	\$15,794	\$0	\$18,685	\$197,719	
11	ACCESSORY ELECTRIC PLANT	\$39,477	\$23,169	\$40,631	\$0	\$0	\$103,277	\$8,979	\$0	\$22,375	\$134,631	
12	INSTRUMENTATION & CONTROL	\$12,562	\$3,797	\$10,536	\$0	\$0	\$26,894	\$2,414	\$1,345	\$5,493	\$36,146	
13	IMPROVEMENTS TO SITE	\$5,886	\$2,892	\$15,404	\$0	\$0	\$24,182	\$2,388	\$0	\$7,971	\$34,541	
14	BUILDINGS & STRUCTURES	\$0	\$7,904	\$10,958	\$0	\$0	\$18,862	\$1,705	\$0	\$3,335	\$23,902	
CALCULATED TOTAL COST		\$2,507,989	\$204,847	\$868,455	\$0	\$0	\$3,581,291	\$338,447	\$180,304	\$702,016	\$4,802,057	

Table 3-16
Case 3a Reference Plant – Total Overnight Cost Breakdown

Owner's Costs	\$ x \$1,000
Preproduction Costs	
6 months All Labor	\$33,441
1 Month Maintenance Materials	\$6,427
1 Month Non-Fuel Consumables	\$1,007
1 Month Waste Disposal	\$1,259
25% of 1 Months Fuel Cost at 100% CF	\$7,178
2% of TPC	\$96,041
Total	\$145,354
Inventory Capital	
60 day supply of fuel at 100% CF	\$56,640
60 day supply of non-fuel consumables at 100% CF	\$1,401
0.5% of TPC (spare parts)	\$24,010
Total	\$82,051
Initial Cost for Catalyst and Chemicals	\$11,284
Land	\$900
Other Owner's Cost	\$720,309
Financing Costs	\$129,656
Total Owner's Costs	\$1,089,553
Total Overnight Costs (TOC)	\$5,891,610

3.8 OPERATING COSTS

Table 3-17 shows the operating expenditure (OPEX) breakdown for the Case 3a reference plant.

Table 3-17
Case 3a Reference Plant – Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Case 3a: Shell CTM w/Rectisol®-based AGR and CO2 Capture				
Plant Size (mtpd Methanol)	10,100				
Primary/Secondary Fuel:	PRB/Natural Gas				
Design/Construction	5 years		Book Life (yrs):	20	
TPC (Plant Cost) Year	June 2011		TPI Year:	2016	
Capacity Factor (%)	90		CO2 Captured (TPD)	19380	
OPERATING & MAINTENANCE LABOR					
Operating Labor					
Operating Labor Rate (base):	\$39.70 \$/hr				
Operating Labor Burden:	30.00 % of base				
Labor Overhead Charge	25.00 % of labor				
Operating Labor Requirements per Shift	units/mod		Total Plant		
Skilled Operator	2.0		2.0		
Operator	10.0		10.0		
Foreman	1.0		1.0		
Lab Tech's etc	3.0		3.0		
TOTAL Operating Jobs	16.0		16.0		
			Annual Cost		
			\$		
Annual Operating Labor Cost			\$7,233,658		
Maintenance Labor Cost			\$46,271,978		
Administration & Support Labor			\$13,376,409		
Property Taxes and Insurance			\$96,041,131		
TOTAL FIXED OPERATING COSTS			\$162,923,176		
VARIABLE OPERATING COSTS					
Maintenance Material Cost			\$69,407,968		
Consumables	Consumption	Unit	Initial Fill		
	Initial	/Day	Cost	Cost	
Water(/1000 gallons)	0	6,902	1.67	\$0	\$3,795,083
Chemicals					
MU & WT Chem (lb)	0	41122	0.27	\$0	\$3,618,240
Carbon (Hg Removal) (lb)	263903	402	1.63	\$430,162	\$215,253
MEA Solvent (ton)	152	0.21	3751.70	\$569,647	\$255,903
Water Gas Shift Catalyst (ft3)	10514	7.20	771.99	\$8,116,617	\$1,826,896
Methanol Synthesis Catalyst	4054	3.70	534.68	\$2,167,532	\$649,538
Claus Catalyst (ft3)	w/equip	2.05	203.15	\$0	\$136,916
Subtotal Chemicals			\$11,283,957	\$6,702,747	
Other					
Supplemental Electricity (MWh consumed)	0	18.5	62.33	\$0	\$378,778
Gases, N2 etc. (/100scf)	0	0	0.00	\$0	\$0
LP Steam (/1000 lbs)	0	0	0.00	\$0	\$0
Subtotal Other			\$0	\$378,778	
Waste Disposal:					
Spent Mercury Catalyst (lb)	0	402	0.65	\$0	\$85,837
Flyash (ton)	0	0	0.00	\$0	\$0
Slag (ton)	0	1638	25.11	\$0	\$13,515,321
Subtotal Waste Disposal			\$0	\$13,601,158	
By-products & Emissions					
Sulfur (tons)	0	138.3	0.00	\$0	\$0
Supplemental Electricity (MWh generated)	0	0	-59.59	\$0	\$0
Subtotal By-Products			\$0	\$0	
TOTAL VARIABLE OPERATING COSTS			\$11,283,957	\$93,885,734	
Coal (tons)	0	19,418	36.57	\$0	\$233,276,555
Natural Gas (MMBtu)	0	38,152	6.13	\$0	\$76,827,051

3.9 METHANOL PRODUCT REQUIRED SELLING PRICE

Table 3-18 shows a summary of the power output, CAPEX, OPEX, and methanol product RSP for the Case 3a reference plant. The Case 3a reference plant methanol RSP is estimated to be \$1.46/gal under the loan guarantee finance structure and \$1.72/gal under the commercial fuels finance structure.

Table 3-18
Case 3a CTM Plant – Overall Performance and Economic Summary

Case	Case 3a	
CAPEX, \$MM		
Total Installed Cost (TIC)	\$3,581	
Total Plant Cost (TPC)	\$4,802	
Total Overnight Cost (TOC)	\$5,892	
OPEX, \$MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC _{fix})	\$162.9	
Variable Operating Cost, less Fuel (OC _{var})	\$103.9	
Coal Feedstock (OC _{coal})	\$259.2	
Natural Gas Feedstock (OC _{NG})	\$85.4	
Import/(Export) Power (OC _{Power})	\$0.4	
Total OPEX	\$611.8	
Plant Output		
Crude Methanol Product, tons per year	4,108,017	
Net Power Output, MWe	-0.77	
Required Selling Price^A		
Excluding CO₂ TS&M^{BE}, \$/ton	424.2	500.7
Including CO₂ TS&M^{BE}, \$/ton	441.5	517.9
RSP Component Details (\$/gal)		
Capital ^B	0.90	1.15
Fixed O&M	0.15	
Variable O&M	0.08	
Coal	0.21	
Natural Gas	0.07	
Power	0.00	
CO ₂ TS&M	0.06	
RSP^B Total (\$/gal)	1.46	1.72
Costs of CO₂ Captured^{B,C} (\$/tonne)	18.8	20.9
Costs of CO₂ Avoided^{B,D} (\$/tonne)	29.8	31.9

^A Capacity factor assumed to be 90 percent

^B Values shown are for two financial structures

The first (lower value) is based on the loan guarantee finance structure

The second (higher value) is based on the commercial fuels finance structure

^C Excludes CO₂ TS&M

^D Includes CO₂ TS&M

^E Based on 301.73 gallons/ton

4.1 PROCESS OVERVIEW

The Case 3b AR Compact Gasifier with RTI WDP CTM plant, like the Case 3a reference plant, uses Montana PRB coal and is designed to generate enough syngas to produce a nominal 10,000 mtpd of methanol. The Case 3b plant is equipped with gas and steam turbines to generate power via NGCC to meet the plant's auxiliary power demands. No power import is required, and excess power is exported to the grid as byproduct for additional revenue. The Case 3b plant is designed to capture CO₂ with a carbon capture efficiency of more than 90% of the carbon in the coal and natural gas.

The Case 3b plant is assumed to operate with an annual on-stream CF of 90% or 7,884 hr/year at full capacity.

The Case 3b plant has the following characteristics that differentiate it from the Case 3a reference plant:

- The AR DSP feed system replaces the reference case's lockhopper system used to feed dried coal to the gasifier. Nexant had previously evaluated the AR DSP feed system in comparison to the Shell lockhopper feed system in a separate study (DE-FE0012062) for DOE on behalf of AR. The results from this earlier study has been used to establish the performance and cost of the DSP feed system for this case.
- The AR compact gasifier replaces the Shell gasifier from Case 3a.
- Syngas leaving the gasifier that has been quenched enters RTI's WDP for sulfur removal at above 760°F instead of going through a low temperature scrubber as in Case 3a.
- After sulfur removal in RTI's WDP, the treated syngas in Case 3b then enters the WGS process. Instead of a sour shift WGS process used in Case 3a, the Case 3b AFWGS process consists of fixed-bed reactors (using commercial high-temperature sweet WGS catalyst) combined in such a manner as to significantly reduce the overall steam consumption and reactor capital cost while still meeting the catalyst manufacturer's steam to CO recommendations. These reactors are operated at standard temperatures for commercial high-temperature sweet WGS processes, but at a higher inlet temperature than commercial sour WGS process.
- After the hydrogen-rich shifted syngas is cooled, it enters an AACRP unit for CO₂ capture. Unlike Rectisol[®], which uses a physical solvent, AACRP uses activated methyldiethanolamine, which is a chemical solvent. As ~99.9% of the sulfur compounds have been removed upstream by the WDP process, the AACRP process only has to remove CO₂ and is less complicated than the two-stage Rectisol[®] process. The AACRP unit also captures ~99% of any residual sulfur left in the syngas following WDP along with the CO₂, resulting in an overall system total sulfur reduction of >99.99% (sub-ppmv total sulfur in the final cleaned syngas).

-
- The DSRP replaces the Claus process used in Case 3a. In the DSRP, sulfur leaving the WDP process in the form of SO₂ is reduced by a slipstream of shifted, hydrogen-rich syngas, forming elemental sulfur, H₂S and COS. The elemental sulfur is condensed while the remaining H₂S and COS are re-oxidized in the presence of air to SO₂. The SO₂ is then removed in a lime scrubber, forming gypsum (CaSO₄·2H₂O). This combined approach for sulfur capture results in very low net SO₂ emissions.

Due to the different cold gas efficiency of the AR compact gasifier, the AR compact gasifier will consume a different amount of coal feed when compared with the Shell gasifier in Case 3a in order to produce the same amount of syngas to produce nominally 10,000 mtpd of methanol.

4.2 CASE 3B PROCESS DESCRIPTION

Case 3b is modeled by adapting the Case 3a reference model with gasifier and syngas cleanup process information provided by AR and RTI. The system description below follows the BFD in Figure 4-3 and stream numbers referenced in the same figure. The overall BFD for the Case 3b plant is shown in Figure 4-3, with the accompanying stream flows shown in Table 4-1.

Additional descriptions of the Case 3b plant's various processes are provided below. To better visualize the different unit operations in the Case 3b plant, simplified PFDs of the various plant processes are depicted in Figures 4-3 through 4-10. Tables 4-2 through 4-9 provide the model-generated process data for the numbered streams referenced in the PFDs.

Coal Preparation and Drying

Same as Case 3a except no supplemental natural gas firing is required in the dryer. Figure 4-3 depicts the coal drying process while Table 4-2 presents the mass balances.

AR Dry Solids Pump (DSP) Coal Feed System

Dried coal from the atmospheric storage silo enters the AR DSPs via gravity flow. Three DSPs, each with a nominal capacity of 1,000 tons per day (tpd), are required to service each of the gasification trains. The DSPs increase the pressure of the coal from atmospheric to 1,100 psia and subsequently discharge the coal continuously to a pressurized feed bin.

Coal is continuously withdrawn from the pressurized feed bin and conveyed by HP CO₂ via a single feed line to each gasifier. To maximize conversion efficiency of fuel to syngas in the gasifier, AR uses its proprietary static splitter system.

Air Separation Unit

Same as Case 3a except no oxygen is routed to the Claus plant.

AR Compact Gasifier

For the Case 3b plant, five AR compact gasifier trains operating at 915 psia are needed to generate the required amount of syngas. Each gasification train includes a single gasifier with a nominal capacity of 3,000 tpd sized for a bituminous (Illinois #6) coal feed. Based on recent pilot

plant gasifier testing under DE-FE0023577 with PRB coal, demonstrating >98% conversion in less than 1/3 the residence time associated with the nominal design gasifier capacity, each of these gasifiers is estimated to provide a minimum of 3,600 tpd capacity operating on the reference sub-bituminous coal feedstock. This allows four out of five gasifiers to gasify 14,400 tpd of coal out of the 14,500 required for Case 3b, providing 99% capacity factor without an installed spare.

Fuel feeds from the pressurized feed bin via the dense phase feed line, conveyed by HP CO₂. To maximize conversion efficiency of fuel to syngas, AR splits the feedstock from a single feed line into many small injection ports via its proprietary static splitter system. The injection ports maximize mixing of coal and oxygen to initiate the gasification reaction.

The AR compact gasifier is oriented in a vertical, down-firing position. The gasifier reaction is initiated with a torch burner, which is ignited at full gasifier pressure. The ignition torch runs on natural gas and oxygen.

The gasifier injector faceplate and the gasifier liner are water cooled to maintain the metal components at temperatures conducive to long life. The cooling water needs to be clean enough (HP BFW quality) to prevent scale buildup or clogging of internal cooling passages.

The gasifier's raw syngas product is partially quenched from about 2,350°F to around 800°F through the introduction of quench water spray. The quench water enters the gasifier through multiple hydraulic atomizing spray nozzles.

The solids are removed as slag and ash. Liquid slag is solidified in a water bath and removed via a lock hopper system. Ash carried over with the syngas is removed in a candle filter. The ash is also removed via a lock hopper system.

The AR DSP, ASU, gasification and quench processes are shown in Figure 4-4 and their mass balances are presented in Table 4-3.

Dry Solids Removal

Same process description as Case 3a but operating at a higher pressure. The solids removal process is shown in Figure 4-4.

After solids removal, the syngas then goes through a raw gas cooler, which lowers the gas temperature and contributes to the production of HP steam for use in steam cycle. The syngas then enters the RTI WDP unit immediately downstream for sulfur removal.

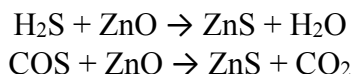
Unlike Case 3a, the syngas scrubber is eliminated in Case 3b as the hot syngas at around 760°F enters the RTI WDP directly without quench cooling.

Warm Syngas Cleanup Process

RTI's advanced warm syngas cleanup process consists of five major system components: RTI WDP, RTI DSRP, AFWGS, Low Temperature Gas Cooling (LTGC) and the AACRP unit. These are described in greater detail below.

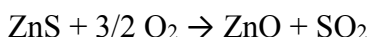
Warm Desulfurization Process

The WDP process, shown in Figure 4-2, uses transport-bed reactors that are similar to commercial Fluid Catalytic Cracking (FCC) reactor designs. It consists of a pair of these reactors: an adsorber and a regenerator. Hot syngas leaving the candle filter is routed to the WDP adsorber where it is contacted with circulating Zn-containing attrition-resistant sorbent (developed and patented by RTI) to remove the sulfur-bearing compounds, in the form of H₂S and COS, from the syngas. The following reactions take place when the sorbent contacts the raw syngas:



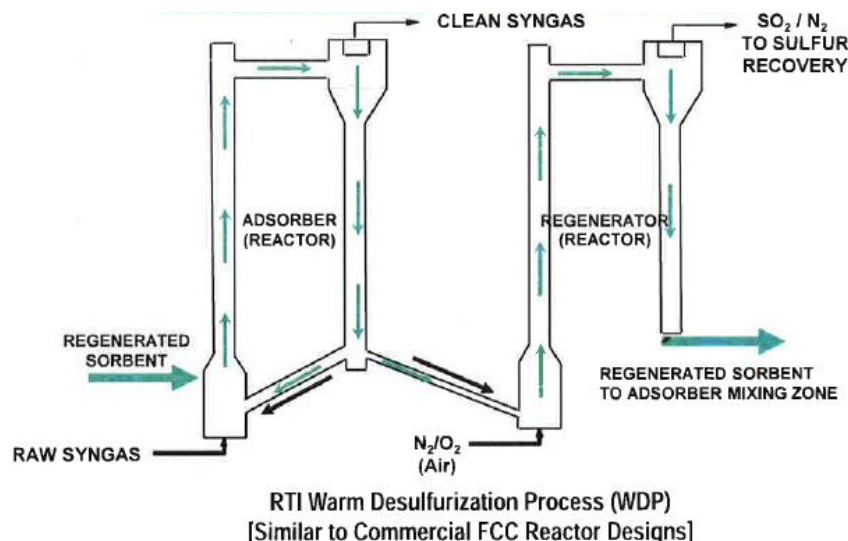
Regenerated sorbent from the regenerator, along with recycled sorbent from the adsorber standpipe, contact the raw syngas, which enters the adsorber near the bottom of the unit. The treated, essentially sulfur-free syngas is separated from sorbent via a cyclone. Any remnant attrited or fine particulate solids entrained in the essentially sulfur-free syngas are removed in a filter. A majority of the sorbent separated by the cyclone is recycled to the adsorber via a standpipe, while a portion of the sorbent is fed to the regenerator.

Within the regenerator, oxidation of the ZnS containing sorbent takes place, producing SO₂ and regenerating ZnO, per the following reaction:



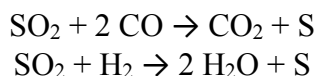
The regenerator uses air as the oxidant. Air is compressed in a multi-stage air compressor up to the regenerator operating pressure before it is fed into the regenerator. The oxidation reaction is exothermic, raising the temperature of the resulting mixture. The regenerator offgas containing SO₂ is heat exchanged with the compressed air before the offgas enters the DSRP. The regenerated sorbent is recycled back to the adsorber, where it adsorbs H₂S and COS again.

Figure 4-1
WDP Process Schematic



Direct Sulfur Reduction Process

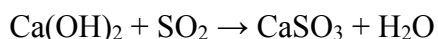
The offgas from the regenerator contains essentially SO_2 and N_2 . It goes through a filter to remove any entrained solids and is cooled before it is sent into the fixed-bed DSRP reactor where SO_2 is reduced to elemental sulfur according to the following reactions:

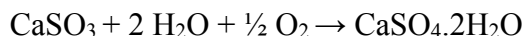


The reducing gas is provided by a hydrogen-rich syngas slip stream from downstream of the shift reactors. A slight excess of the reducing gas is used to ensure complete reduction of the SO_2 . Some H_2S and COS are formed alongside the elemental sulfur. The product stream from the DSRP reactor is sent on to a sulfur condenser unit where the elemental sulfur is condensed and separated. Heat is recovered in the condenser by making low pressure steam.

The condenser overhead gas still contains some residual H_2S and COS . These are re-oxidized to SO_2 in a fixed-bed oxidation reactor containing a redox catalyst in the presence of compressed air, which functions as the oxidant. The compressed air is a slipstream drawn from the air compressor in the WDP section.

Finally, the SO_2 -containing gas leaving the oxidation reactor is cooled and sent to a lime scrubber downstream. Lime, or calcium hydroxide ($\text{Ca}(\text{OH})_2$), reacts with the SO_2 in the presence of oxygen to form gypsum ($\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$) per the following reaction, which is akin to the Wet Flue Gas Desulfurization (WFGD) process used for scrubbing flue gas.





The gypsum leaves the bottom of the scrubber as a byproduct of the CTM plant. The treated, sulfur-free overhead gas is split into three portions. One portion is recycled to the DSRP to help control reactor temperature rise due to the exothermic heat from the DSRP reactions. The second portion is compressed and routed to the WDP process to be utilized as stripping or fluidizing gas in the WDP adsorber and regenerator. The remaining portion is vented to the atmosphere.

Advanced Fixed-Bed Water Gas Shift

The treated syngas from the WDP enters the AFWGS process. In order to achieve a 2:1 ratio of H₂ to CO in the final syngas to the methanol synthesis reactors, approximately 64 percent of the coal-derived syngas is shifted, with the remainder bypassing the shift reactors. Part of the steam for the WGS reaction is provided partly by vaporized quench water in the AR compact gasifier. IN RTI's AFWGS process a series of fixed-bed reactors using a commercial high-temperature sweet WGS catalyst are combined in such a manner that enables a significant reduction in the overall steam consumption and capital cost of the AFWGS process while still meeting the catalyst manufacturer's steam to CO recommendations and achieving the high CO conversion required by the process.

The RTI WDP, DSRP and AFWGS processes are shown in Figure 4-5 and the mass balances are presented in Table 4-4.

Low Temperature Gas Cooling and Mercury Removal

The shifted syngas goes through a series of heat exchangers to generate various levels of steam and preheat BFW before it finally undergoes cooling by cooling water. The cooled syngas is sent to a knockout (KO) drum where the condensate is drained. Mercury is removed from the overhead syngas leaving the KO drum via the process described in Section 3.2.4. RTI has also developed a warm-gas mercury capture process that operates at ~350°F-390°F, though this was not used here. The LTGC and mercury removal processes are shown in Figure 4-6 and the mass balances presented in Table 4-5.

Activated Amine CO₂ Removal Process

The AACRP is based on the activated methyldiethanolamine process, which is marketed commercially by companies such as BASF, Shell and UOP for the removal of acid gases like H₂S and CO₂. The scrubbing agent is an aqueous alkaline amine solution. For this application, the CO₂-containing syngas is passed through an absorber that contains a circulating alkaline amine scrubbing solution, where the bulk of the CO₂ removal takes place. An acid-base reaction occurs where the CO₂ reacts with alkaline amine and is captured in solution.

To achieve the desired extent of CO₂ capture, and to ensure that the product gas meets the volatile organic compound (VOC) specifications, a lean amine absorber is used to treat the syngas leaving the main alkaline amine absorber. Water enters at the top of this absorber where it contacts and scrubs the CO₂-lean syngas. This serves to remove any entrained alkaline amine.

The rich, CO₂-loaded solution is sent to a regenerator to release the absorbed CO₂. The solution is first depressurized, flashing off some CO₂ in the process and helping to reduce the overall heat of CO₂ regeneration. The low-pressure solution is then sent to a thermal regenerator, where heat is applied to release the remaining CO₂. Regenerated alkaline amine solution is recycled to the absorber and used again.

For the AACRP, the CO concentration in the CO₂ product is about 700 ppmv. Although this is higher than the CO limit (35 ppmv) for saline reservoir CO₂ sequestration as shown in Table 2-5, it is still within the range stated in literature (10-5,000 ppmv). It should also be noted that the CO concentration of the AACRP CO₂ product is one order of magnitude lower than in Case 3a as AACRP is more selective towards absorbing CO₂ when compared with the Rectisol[®] process.

CO₂ Compression and Dehydration

CO₂ from the AACRP is generated at a single pressure of 20.7 psia. The CO₂ stream is compressed to a supercritical condition at 2,215 psia using a multiple-stage, intercooled compressor. During compression, the CO₂ stream is dehydrated to a dew point of -40°F using a thermal swing adsorptive dryer. The dehydrated CO₂ is transported to the plant fence line for sequestration outside the battery limit (OSBL).

The AACRP and CO₂ compression process are shown in Figure 4-7, while the material balance is shown in Table 4-6. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table.

Methanol Reactor and Synthesis Loop

The methanol synthesis process is shown in Figure 4-8, while the material balance is shown in Table 4-7. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table. The process follows the description stated earlier in Section 3.2.4.

For the Case 3b plant, the syngas feed compressor to the methanol synthesis reactors has been eliminated due to the higher system pressure upstream. Also, there is excess purge gas after taking into account the fuel requirements for coal drying. This excess gas is routed to the NGCC to produce power, thus cutting back on the natural gas firing rate for power generation.

NGCC

Same as Case 3a. The PFD for the NGCC section is shown in Figure 4-9, while the material balance is shown in Table 4-8. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table.

Power Cycle Flue Gas Post Combustion CO₂ Capture and Compression

Like in Case 3a, Fluor's Econamine FG+ process was used to capture CO₂ from the flue gas leaving the NGCC HRSG. The CO₂ leaving the Econamine capture process, like the AACRP CO₂ product, still contains moisture. The two moisture-bearing CO₂ streams are thus combined

and compressed to 2,215 psia by a multi-stage, intercooled centrifugal compressor that is equipped with a thermal swing adsorptive dryer, which dehydrates the CO₂ stream to a dew point of -40°F. The virtually moisture-free supercritical CO₂ stream is delivered to the plant B/L as sequestration ready.

Figure 4-2
Case 3b Plant – Overall BFD

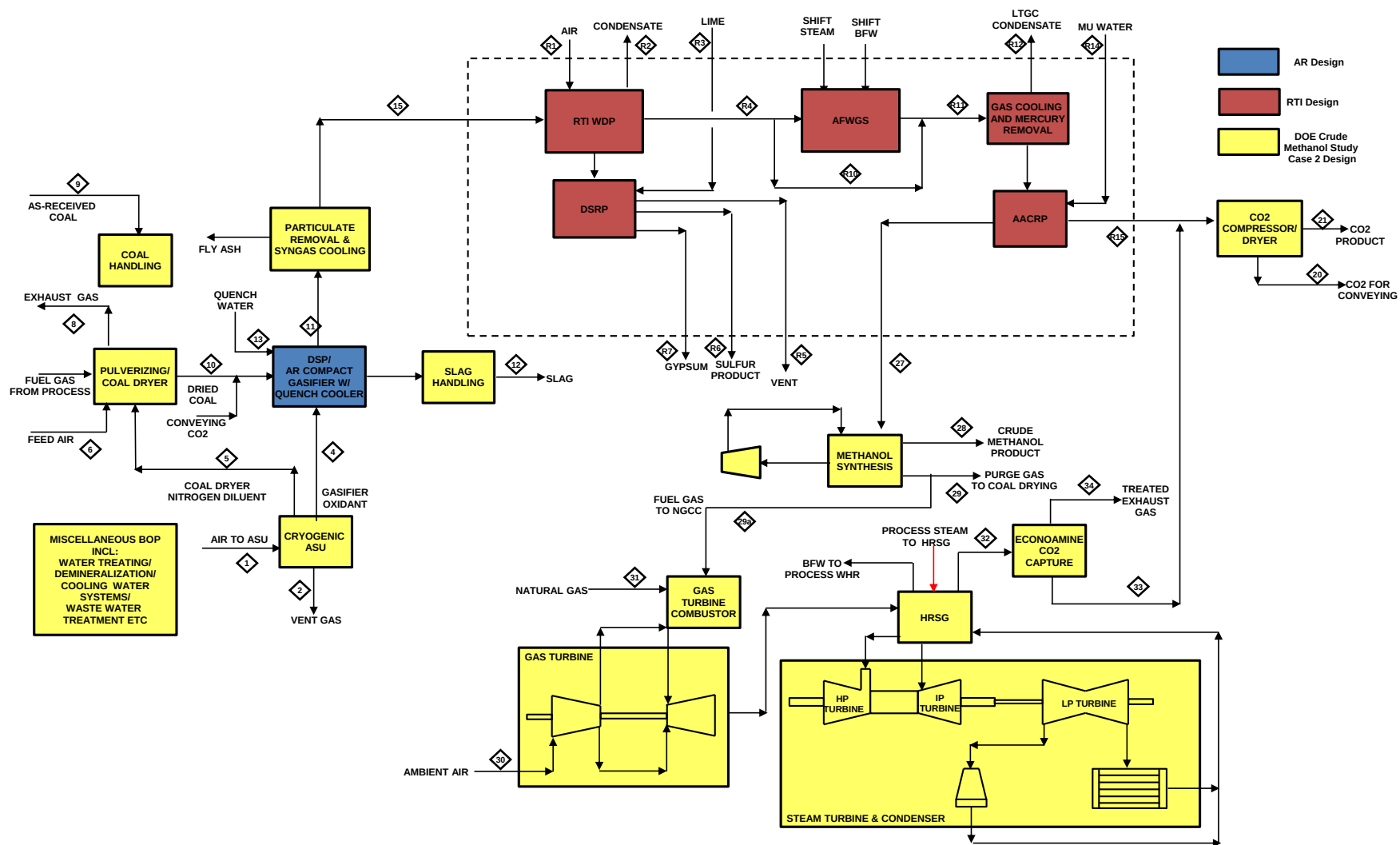


Table 4-1
Case 3b Plant – Overall Stream Table

Description	Air to ASU	ASU Vent Gas	Gasifier Oxidant	Nitrogen Diluent to Dryer	Air Feed to Coal Dryer	Coal Dryer Exhaust	As-Received Coal	Dried Coal	Quenched Syngas	Slag	Quench Water	Syngas to RTI WDP
Stream No.	1	2	4	5	6	8	9	10	11	12	13	15
V-L Mole Fraction												
Ar	0.0092	0.0219	0.0300	0.0023	0.0092	0.0077	0	0	0.0053	0	0	0.0053
CH4	0	0	0	0	0	0	0	0	0.0023	0	0	0.0023
CO	0	0	0	0	0	0	0	0	0.4251	0	0	0.4251
CO2	0.0003	0.0052	0	0	0.0003	0	0	0	0.0134	0	0	0.0134
COS	0	0	0	0	0	0	0	0	0.0002	0	0	0.0002
H2	0	0	0	0	0	0	0	0	0.1719	0	0	0.1719
H2O	0.0099	0.1535	0	0	0.0099	0.1567	0	0	0.3732	0	1.0000	0.3732
H2S	0	0	0	0	0	0	0	0	0.0021	0	0	0.0021
N2	0.7732	0.6989	0.0200	0.9921	0.7732	0.7947	0	0	0.0057	0	0	0.0057
NH3	0	0	0	0	0	0	0	0	0.0007	0	0	0
O2	0.2074	0.1206	0.9500	0.0054	0.2074	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0.0000	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0	0	0	0	0	0
Methanol	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0
Propanol	0	0	0	0	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane	0	0	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0	0	1.0000	0	1.0000	1.0000
V-L Flowrate, lbmol/hr	128485	8141	26484	93860	18222	136198	0	0	151097	0	53860	151097
V-L Flowrate, lb/hr	3707669	222286	851648	2633735	525820	3682889	0	0	3032729	0	970303	3032728
Solids Flow Rate, lb/hr	0	0	0	0	0	0	1528260	1206840	0	125127.4	0	0
Temperature, F	59	64	292	70	59	190	59	190	826		250	755
Pressure, psia	14.7	16.4	1000.0	18.0	14.7	14.3	14.7	14.3	915.0	915.0	980.0	905.0

Table 4-1 (cont'd)
Case 3b Plant – Overall Stream Table

Description	Air to RTI WDP	Comp Air KO Water	Lime	Sweet Syngas	DSRP Vent Gas	Sulfur Product	Gypsum	Bypass Syngas	Syngas to Cooling	Cooled Syngas KO Water	Syngas to aMDEA Unit	Makeup Wash Water	aMDEA CO2 Out
Stream No.	R1	R2	R3	R4	R5	R6	R7	R10	R11	R12	R13	R14	R15
V-L Mole Fraction													
Ar	0.0092	0	0	0.0053	0.0131	0	0	0.0053	0.0041	0	0.0062	0	0
CH4	0	0	0	0.0023	0	0	0	0.0023	0.0018	0	0.0027	0	0
CO	0	0	0	0.4214	0	0	0	0.4214	0.1535	0	0.2321	0	0.0007
CO2	0.0003	0	0	0.0173	0.0627	0	0.0105	0.0173	0.1863	0.0002	0.2815	0	0.9093
COS	0	0	0	0.0000	0	0	0	0.0000	0.0000	0	0	0	0
H2	0	0	0	0.1754	0	0	0	0.1754	0.3088	0	0.4667	0	0.0020
H2O	0.0099	1.0000	0.9737	0.3693	0.0001	0	0.9649	0.3693	0.3386	0.9981	0.0012	1.0000	0.0879
H2S	0	0	0	0.0000	0	0	0	0.0000	0.0000	0	0	0	0
N2	0.7732	0	0	0.0083	0.9240	0	0.0005	0.0083	0.0064	0	0.0097	0	0
NH3	0	0	0	0	0	0	0	0	0	0.0017	0	0	0
O2	0.2074	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	1.0000	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0.0263	0	0	0	0.0010	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0.0227	0	0	0	0	0	0
Methanol	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0
Propanol	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane	0	0	0	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.000	1.000	1.000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	4679	33	8571	153559	3499	130	9514	55604	198242	67089	129107	3556	36778
V-L Flowrate, lb/hr	135022	603	167062	3081007	102062	4167	207924	1115633	3886005	1208828	2635413	64062	1531019
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature, F	59	80	75	685	105	305	172	685	844	95	95	122	122
Pressure, psia	14.7	13.0	13.0	894.0	750.8	14.7	14.7	894.0	869.0	844.0	829.0	814.0	20.7

Table 4-1 (cont'd)
Case 3b Plant – Overall Stream Table

Description	Convey'g CO2	CO2 Product	Treated Syngas to MeOH	Crude Methanol Product	MeOH Synthesis Purge Gas	Excess Purge to NGCC	Air to Gas Turbine	Natural Gas Feed	NGCC Flue Gas Exhaust	CO2 to Comp- ression	Treated NGCC Exhaust
Stream No.	20	21	27	28	29	29a	30	31	32	33	34
V-L Mole Fraction											
Ar	0	0	0.0083	0.0001	0.1064	0.1064	0.0092	0	0.0100	0	0.0108
CH4	0	0	0.0036	0	0.0435	0.0435	0	0.9310	0	0	0
CO	0.0008	0.0007	0.3122	0	0.1328	0.1328	0	0	0	0	0
CO2	0.9969	0.9973	0.0302	0.0108	0.2011	0.2011	0.0003	0.0100	0.0367	0.9885	0.0040
COS	0	0	0	0	0	0	0	0	0	0	0
H2	0.0022	0.0019	0.6276	0	0.3235	0.3235	0	0	0	0	0
H2O	0	0	0.0050	0.0167	0.0001	0.0001	0.0099	0	0.0762	0.0115	0.0333
H2S	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0.0131	0.0001	0.1728	0.1728	0.7732	0.0160	0.7447	0	0.8081
NH3	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0.2074	0	0.1325	0	0.1438
SO2	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0	0	0	0	0
Methanol	0	0	0	0.9713	0.0197	0.0197	0	0	0	0	0
Ethanol	0	0	0	0.0010	0	0	0	0	0	0	0
Propanol	0	0	0	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0.032	0	0	0
Propane	0	0	0	0	0	0	0	0.007	0	0	0
n-Butane	0	0	0	0	0	0	0	0.004	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	2940	34551	94905	28979	6571	1216	114330	3690	119065	3974	109716
V-L Flowrate, lb/hr	129095	1517330	1156519	925862	155384	28746	3299204	63929	3391495	173711	3120951
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0
Temperature, F	270	162	148	119	130	130	59	60	272	69	85
Pressure, psia	1100.0	2215	814	40	717	717	14.5	474.7	14.9	24	15.0

Figure 4-3
Case 3b Plant – Coal Drying BFD

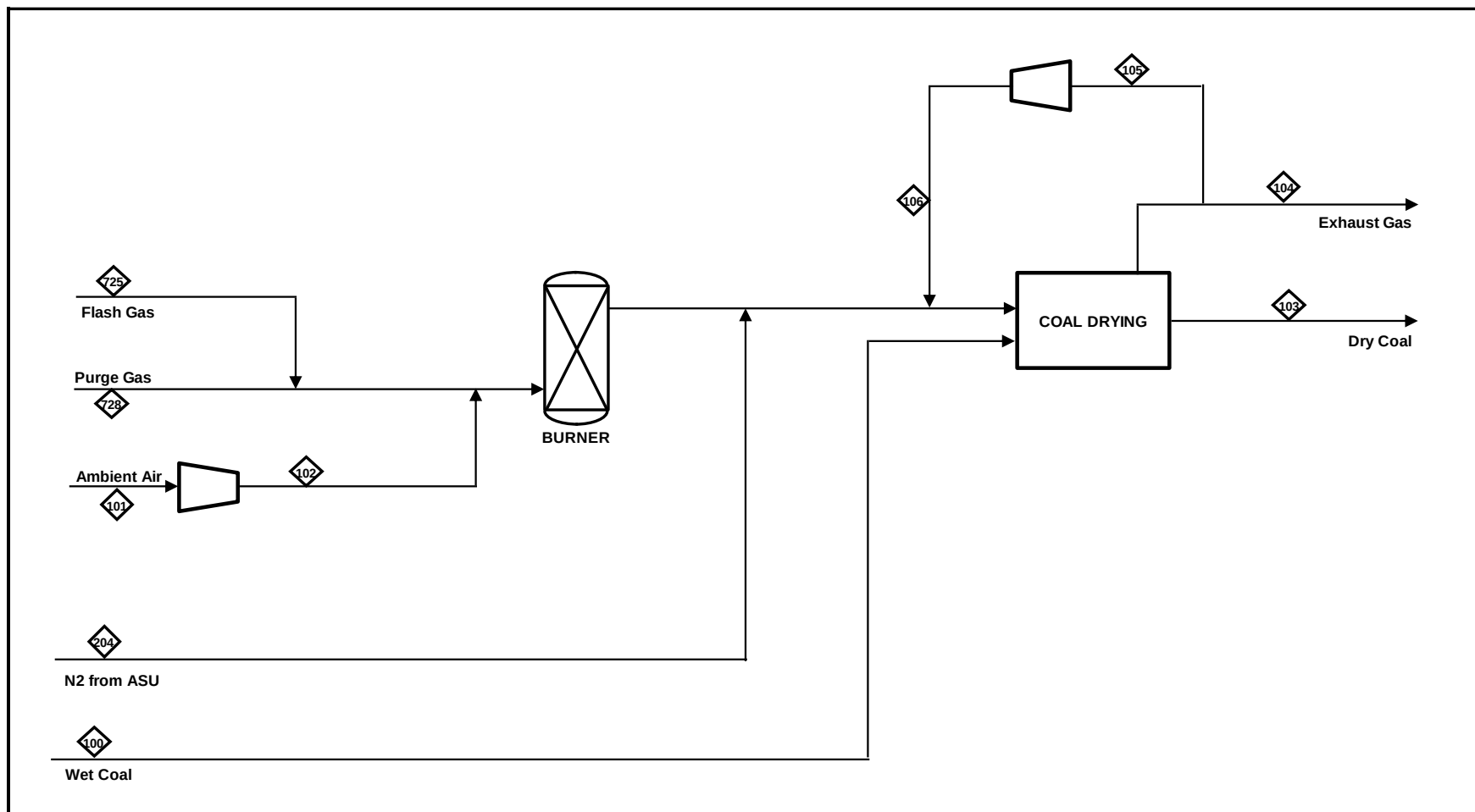


Table 4-2
Case 3b Plant – Coal Drying Stream Table

STREAM	100	101	102	103	104	105	106	204	725	728
Description	Wet Coal	Ambient Air	Feed Air	Dry Coal	Exhaust Gas	LP Recycle Gas	HP Recycle Gas	Dilution Nitrogen	MeOH Flash Gas	MeOH Purge Gas
Mole Flow (Vapor/Liquid)/lbmol/hr										
AR	0	169	169	0	1044	562	562	216	89	570
CH4	0	0	0	0	0	0	0	0	52	233
CO	0	0	0	0	0	0	0	0	38	711
CO2	0	9	9	0	3845	2070	2070	0	1247	1077
COS	0	0	0	0	0	0	0	0	0	0.00268
H2	0	0	0	0	0	0	0	0	39	1733
H2O	0	0	0	0	21156	11392	11392	19	2	1
H2S	0	0	0	0	0	0	0	0	0	0.00114
N2	0	14226	14226	0	108373	58355	58355	93118.15	104	926
NH3	0	0	0	0	0	0	0	0	0	0
O2	0	3817	3817	0	1780	958	958	507	0	0
SO2	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0	370	105
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0.2	0.1
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	0	18222	18222	0	136198	73337	73337	93860	1942	5355
Total V/L Flow, lb/hr	0	527831	527831	0	3684901	1984176	1984176	2633735	75258	126638
Solids Mass Flow, lb/hr	1528260	0	0	1206840	0	0	0	0	0	0
Coal	1528260	0	0	1206840	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	1528260	527831	527831	1206840	3684901	1984176	1984176	2633735	75258	126638
Pressure, psia	14.7	14.7	17	14.3	14.3	14.3	16	18	40	717
Temperature, F		59	90		190	190	219	70	119	130

Figure 4-4
Case 3b Plant – ASU/Coal Gasification PFD

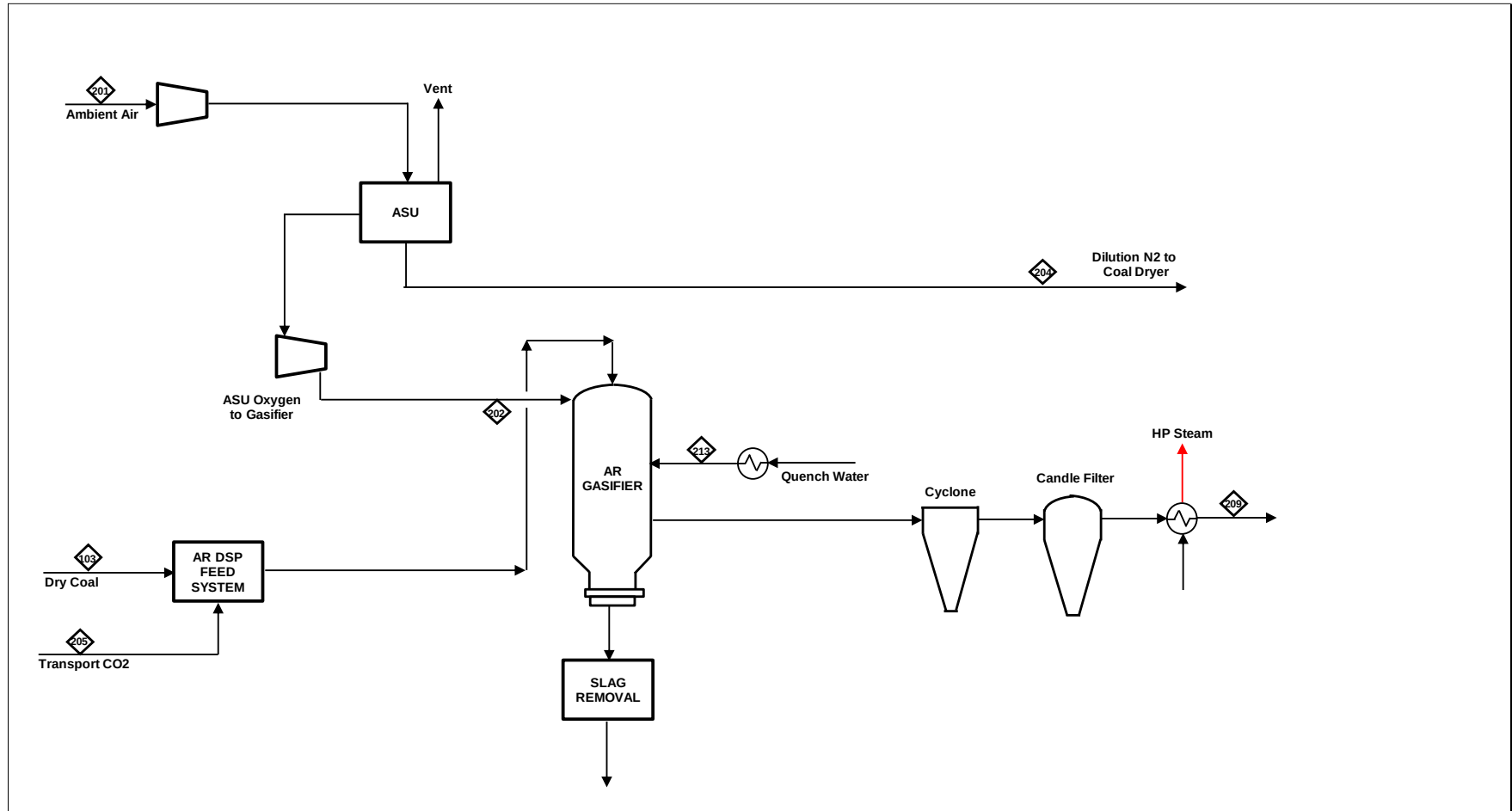


Table 4-3
Case 3b Plant – ASU/Coal Gasification Stream Table

STREAM	201	202	204	205	103	213	214	209
Description	Ambient Air	Oxygen to Gasifier	N2 to Coal Dryer	Conveying CO2	Dry Coal	Quench Water	Slag Out	Raw Cooled Syngas
Mole Flow (Vapor/Liquid)/lbmol/hr								
AR	1188	794	216	0	0	0	0	794
CH4	0	0	0	0	0	0	0	345
CO	0	0	0	2	0	0	0	64237
CO2	42	0	0	2931	0	0	0	2019
COS	0	0	0	0	0	0	0	36
H2	0	0	0	6	0	0	0	25980
H2O	1268	0	19	0	0	53860	0	56396
H2S	0	0	0	0	0	0	0	311
N2	99338	530	93118.15	0	0	0	0	861
NH3	0	0	0	0	0	0	0	111
O2	26648	25159	507	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	3
HCL	0	0	0	0	0	0	0	3
Methanol -- CH3OH	0	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	128485	26484	93860	2940	0	53860	0	151097
Total V/L Flow, lb/hr	3707669	851647.9	2633735	129095	0	970303	0	3032728
Solids Mass Flow, lb/hr	0	0	0	0	1206840	0	125127.4	0
Coal	0	0	0	0	1206840	0	0	0
Ash	0	0	0	0	0	0	125127	0
Carbon	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	3707669	851648	2633735	129095	1206840	970303	125127	3032728
Pressure, psia	14.7	1000	18.0	1100.0	14.3	980	915	905.0
Temperature, F	59	292	70	270		250		755

Figure 4-5
Case 3b Plant – RTI WDP/DSRP/WGS PFD

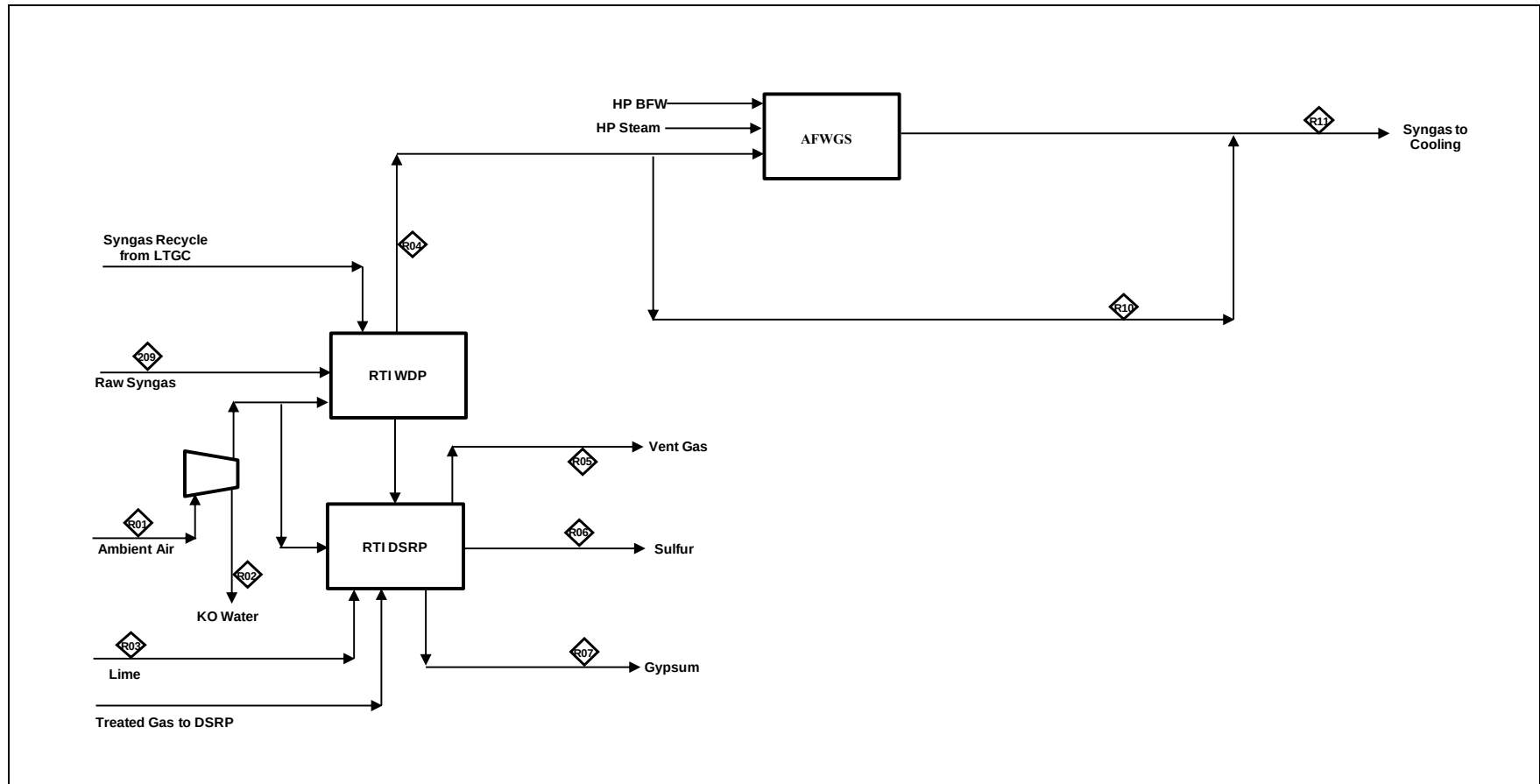


Table 4-4
Case 3b Plant – RTI WDP/DSRP/AFWGS Stream Table

STREAM	209	R01	R02	R03	R04	R05	R06	R07	R10	R11
Description	Raw Syngas	Ambient Air	Comp KO Water	Lime	Treated Gas to Shift	DSRP Vent Gas	Sulfur Product	Gypsum Product	Bypass Syngas	Syngas to LT Cooling
Mole Flow (Vapor/Liquid)/lbmol/hr										
AR	794	43	0	0	812	46	0	0	294	812
CH4	345	0	0	0	351	0	0	0	127	351
CO	64237	0	0	0	64711	0	0	0	23432	30438
CO2	2019	2	0	0	2655	220	0	100	961	36928
COS	36	0	0	0	0	0	0	0	0	0
H2	25980	0	0	0	26935	0	0	0	9753	61208
H2O	56396	46	33	8346	56709	0	0	9179	20534	67119
H2S	311	0	0	0	1	0	0	0	0	1
N2	861	3618	0	0	1274	3233	0	5	461	1274
NH3	111	0	0	0	111	0	0	0	40	111
O2	0	970	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	130	0	0	0
HCN	3	0	0	0	0	0	0	0	0	0
HCL	3	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0
Lime	0	0	0	225	0	0	0	9	0	0
Gypsum	0	0	0	0	0	0	0	216	0	0
Total V/L Flow, lbmol/hr	151097	4679	33	8571	153559	3499	130	9514	55604	198242
Total V/L Flow, lb/hr	3032728	135022	603	167062	3081007	102062	4167	207924	1115633	3886005
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	3032728	135022	603	167062	3081007	102062	4167	207924	1115633	3886005
Pressure, psia	905	14.7	13.0	13.0	894.0	751	14.7	14.7	894	869
Temperature, F	755	59	80	75	685	105	305	172	685	844

Figure 4-6
Case 3b Plant – LTGC PFD

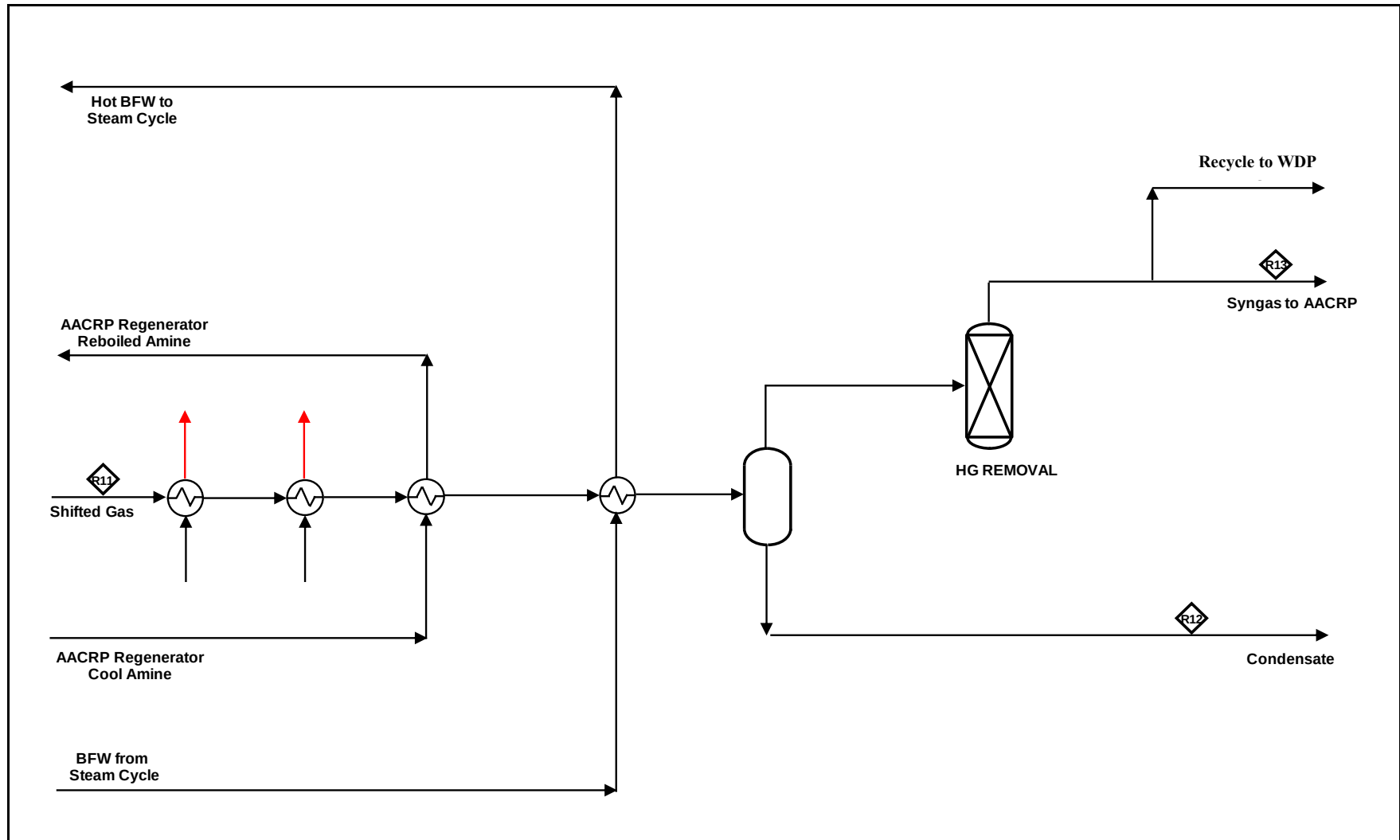


Table 4-5
Case 3b Plant – LTGC Stream Table

STREAM	R11	R12	R13
Description	Shifted Syngas	KO Water	Cooled Syngas
Mole Flow (Vapor/Liquid)/lbmol/hr			
AR	812	0	800
CH4	351	0	345
CO	30438	0	29963
CO2	36928	12	36339
COS	0	0	0
H2	61208	0	60253
H2O	67119	66964	152
H2S	1	0	1
N2	1274	0	1254
NH3	111	111	0
O2	0	0	0
SO2	0	0	0
S	0	0	0
HCN	0	0	0
HCL	0	0	0
Methanol -- CH3OH	0	0	0
Ethanol -- C2H5OH	0	0	0
Propanol -- C3H8O	0	0	0
Ethane -- C2H6	0	0	0
Propane -- C3H8	0	0	0
n-Butane -- C4H10	0	0	0
Total V/L Flow, lbmol/hr	198242	67089	129107
Total V/L Flow, lb/hr	3886005	1208828	2635413
Solids Mass Flow, lb/hr	0	0	0
Coal	0	0	0
Ash	0	0	0
Carbon	0	0	0
Total Flow (Solids + V/L), lb/hr	3886005	1208828	2635413
Pressure, psia	869	844	829
Temperature, F	844	95	95

Figure 4-7
Case 3b Plant – AACRP and CO₂ Compression PFD

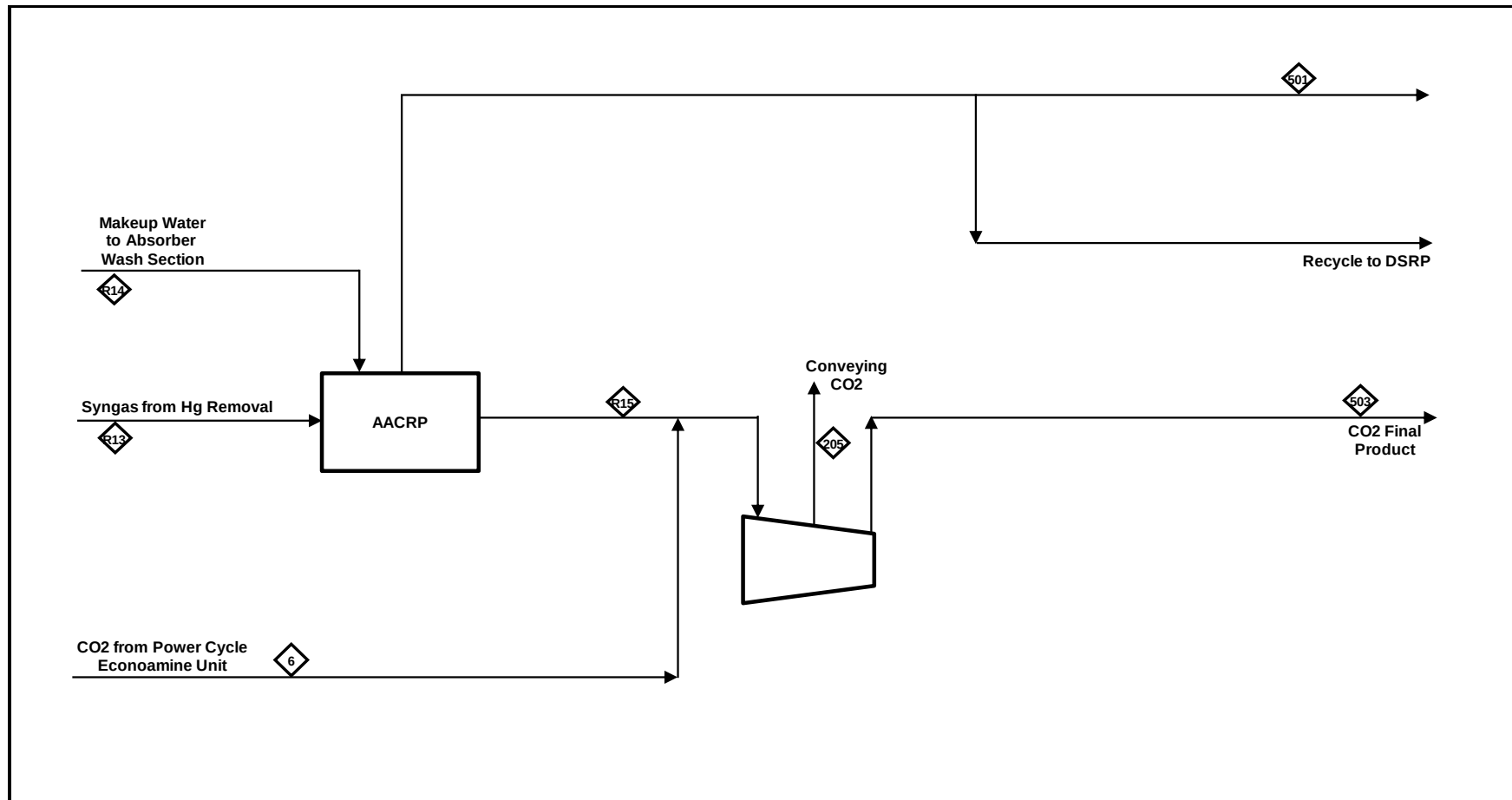


Table 4-6
Case 3b Plant – AACRP and CO₂ Compression Stream Table

STREAM	R13	R14	R15	501	6	205	503
Description	Syngas to AACRP	Makeup Wash Water	AACRP CO2 Product	Syngas to MeOH Synthesis	CO2 from MEA Unit	Convey CO2	CO2 Final Product
Mole Flow (Vapor/Liquid)/lbmol/hr							
AR	812	0	1	790	0	0	1
CH4	351	0	2	340	0	0	2
CO	30438	0	26	29632	0	2	23
CO2	36915	0	33442	2867	3946	2931	34457
COS	0	0	0	0	0	0	0
H2	61208	0	73	59565	0	6	67
H2O	154	3556	3233	471	46	0	0
H2S	1	0	1	0	0	0	1
N2	1274	0	1	1241	0	0	1
NH3	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	131153	3556	36778	94905	3991	2940	34551
Total V/L Flow, lb/hr	2677177	64062	1531019	1156519	174468	129095	1517330
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	2677177	64062	1531019	1156519	174468	129095	1517330
Pressure, psia	829	814	20.7	814	24	1100	2215
Temperature, F	95	122	122	148	69	270	162

Figure 4-8
Case 3b Plant – Methanol Synthesis Plant PFD

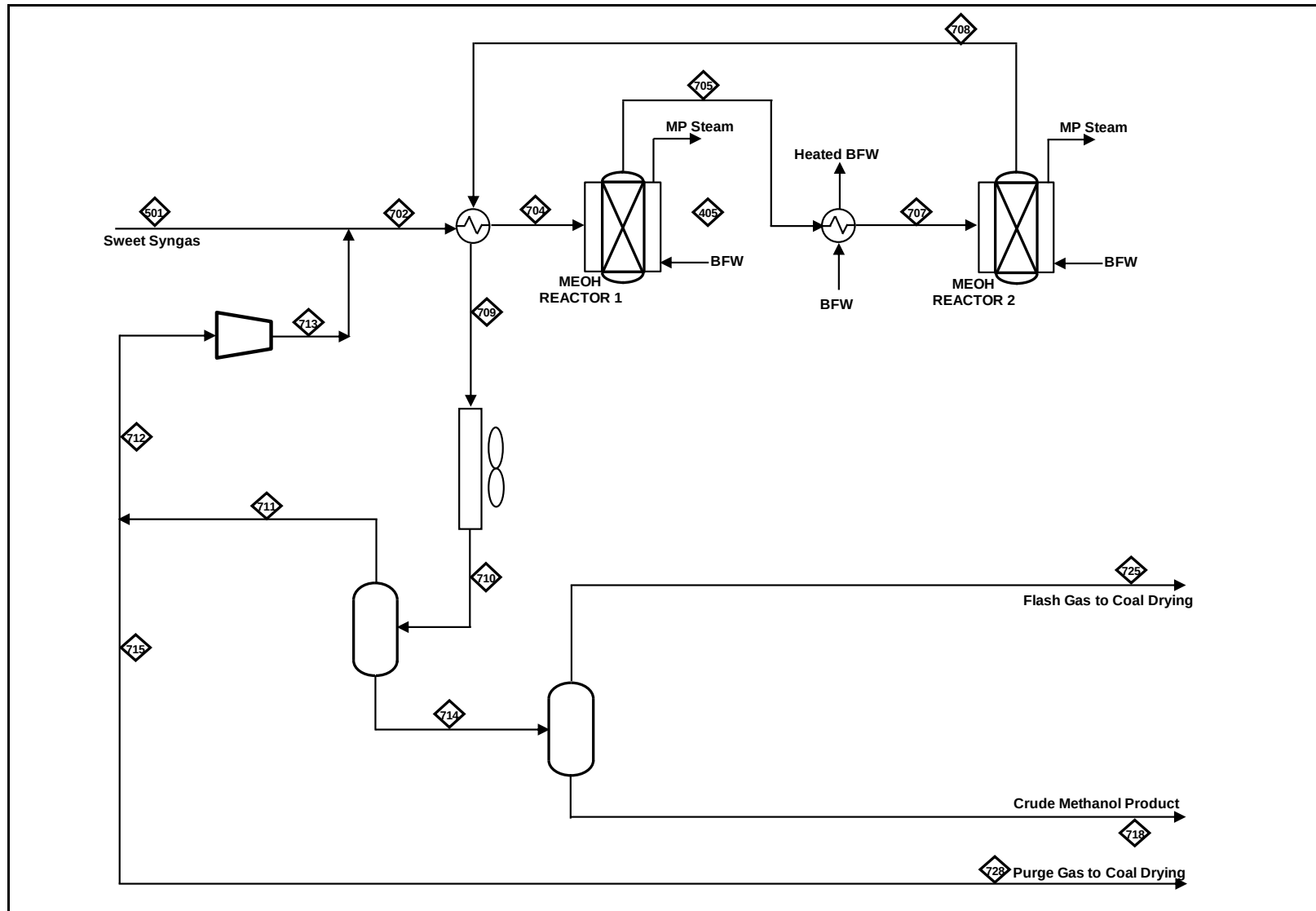


Table 4-7
Case 3b Plant – Methanol Synthesis Plant Stream Table

STREAM	501	704	705	707	708	709	710	711	712	713	714	718	725	728
Description	Reheated Syngas	MEOH Reactor 1 Feed	MEOH Reactor 1 Outlet	MEOH Reactor 2 Feed	MEOH Reactor 2 Outlet	Cooled Raw MeOH	AC Cool Raw MeOH	Flash Gas Ovhd	Recycle Syngas	HP Recycle Syngas	KO Drum Bottoms	Crude MeOH Product	MeOH Flash Gas	MeOH Purge Gas
Mole Flow (Vapor/Liquid)/lbmol/hr														
AR	790	14076	14076	14076	14076	14076	14076	13985	13286	13286	91	2	89	699
CH4	340	5770	5770	5770	5770	5770	5770	5716	5430	5430	54	2	52	286
CO	29632	46213	26363	26363	17492	17492	17492	17453	16582	16582	38	0	38	873
CO2	2867	27980	27793	27793	27995	27995	27995	26434	25112	25112	1561	314	1247	1322
COS	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	59565	99956	59694	59694	42558	42558	42558	42520	40391	40391	39	0	39	2126
H2O	471	485	708	708	500	500	500	15	14	14	485	483	2	1
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	1241	22817	22817	22817	22817	22817	22817	22712	21576	21576	105	2	104	1136
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	2455	22422	22422	31101	31101	31101	2584	2455	2455	28517	28147	370	129
Ethanol -- C2H5OH	0	2	35	35	30	30	30	2	2	2	29	28	0	0
Propanol -- C3H8O	0	0	1	1	1	1	1	0	0	0	1	1	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	94905.3	219753	179679	179679	162341	162341	162341	131420	124848	124848	30920	28979	1942	6571
Total V/L Flow, lb/hr	1156519	4108841	4108806	4108806	4108791	4108791	4108791	3107671	2952322	2952322	1001119	925862	75257.7	155384
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	1156519	4108841	4108806	4108806	4108791	4108791	4108791	3107671	2952322	2952322	1001119	925862	75258	155384
Pressure, psia	814	747	737	732.0	727	722	720	717	717	755	717	40	40	717
Temperature, F	147.8604	400	475	400	430	235	130	130	130	141	130	119	119	130

Figure 4-9
Case 3b Plant -- NGCC PFD

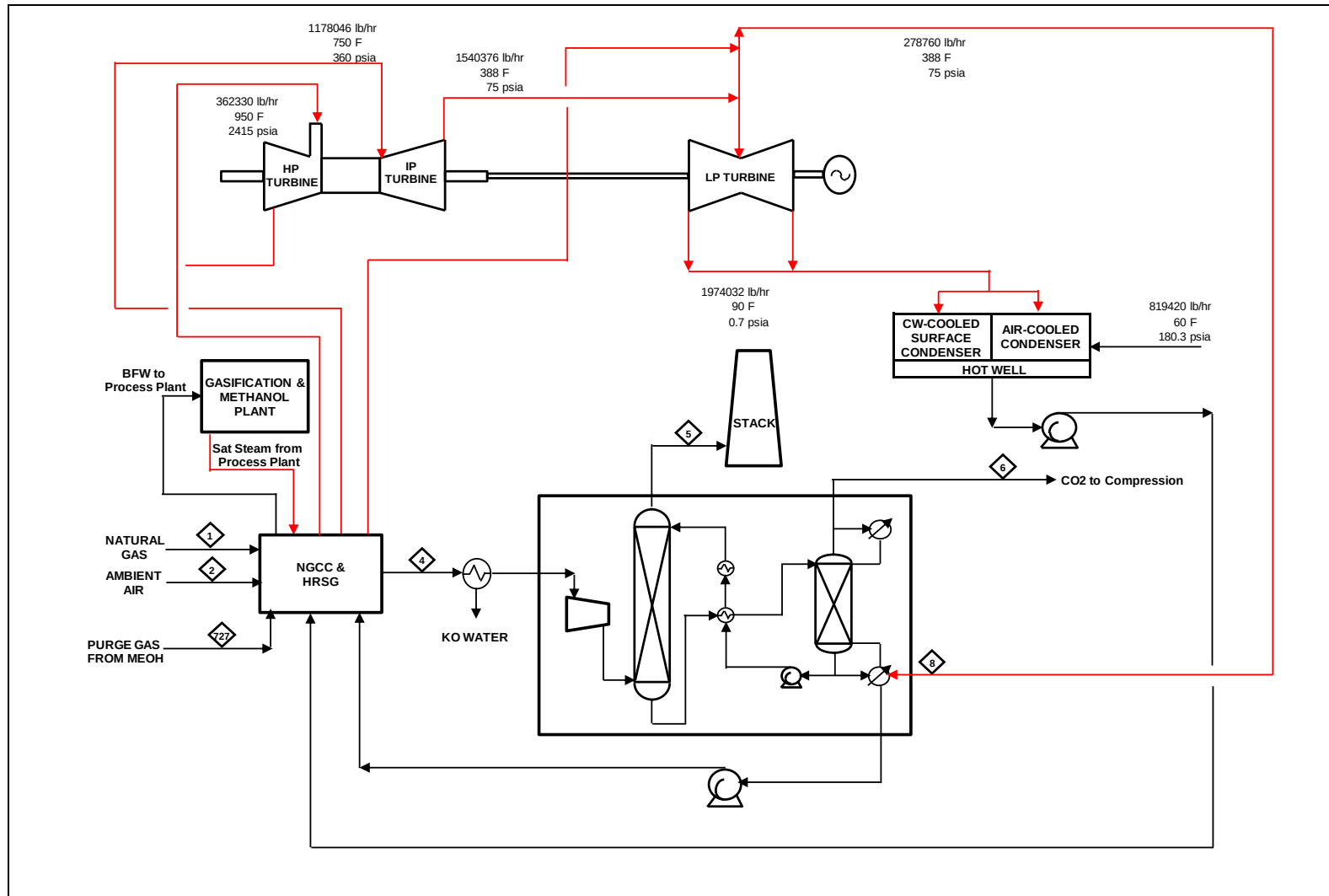


Table 4-8
Case 3b Plant – NGCC Stream Table

STREAM	1	2	4	5	6	8	727
Description	Ambient Air	Natural Gas	Flue Gas to MEA	Treated Gas	CO2 Out	LP Steam	MeOH Purge Gas
Mole Flow (Vapor/Liquid)/lbmol/hr							
AR	1053	0	1182	1182	0	0	129
CH4	0	3435	0	0	0	0	77
CO	0	0	0	0	0	0	161
CO2	57	37	4384	438	3946	0	244
COS	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	393
H2O	1132	0	9081	3661	46	15474	0
H2S	0	0	0	0	0	0	0
N2	88373	59	88643	88643	0	0	210
NH3	0	0	0	0	0	0	0
O2	23715	0	15775	15775	0	0	0
SO2	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0
Ethane -- C2H6	0	118	0	0	0	0	0
Propane -- C3H8	0	26	0	0	0	0	0
n-Butane -- C4H10	0	15	0	0	0	0	0
Total V/L Flow, lbmol/hr	114330	3690	119065	109699	3991	15474	1215
Total V/L Flow, lb/hr	3299430	63929	3391721	3120423	174468	278760	28351
Total Flow (Solids + V/L), lb/hr	3299430	63929	3391721	3120423	174468	278760	28351
Pressure, psia	15	475	15	15	24	260	176
Temperature, F	59	60	272	85	69	35	365

4.3 SPARING PHILOSOPHY

The sparing philosophy for the AR and RTI-provided equipment for the Case 3b CTM plant is provided below. The design has:

- Five trains of gasification, including dry feed system, AR compact gasifier, cyclone and candle filter (5 x 25%)
- One WDP train (1 x 100%)
- One DSRP train (1 x 100%)
- Two AFWGS trains (2 x 50%)
- Two AACRP trains (2 x 50%)

4.4 PERFORMANCE RESULTS

The Nexant-modeled Case 3b plant with CO₂ capture consumes 18,339 tpd of PRB coal and 33.6 MMSCFD of natural gas at the Midwestern site and produces 925,862 lb/hr (10,079 mtpd) of crude methanol. Overall performance for the Case 3b plant is summarized in Table 4-9, which includes auxiliary power requirements. Loads that are unique to the AR gasification system, RTI WDP, DSRP, and AACRP are shown in bold and italicized.

Table 4-9
Case 3b Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case 3b
Gas Turbine Power	127,856
Steam Turbine Power	239,204
TOTAL POWER, kWe	367,060
CTM Plant Auxiliary Load Summary, kWe	
Coal Handling	8,585
Slag Handling	2,178
Incinerator Air Blower	4,521
Incinerator Recycle Blower	1,162
Air Separation Unit	152,581
AR DSP	8,483
Quench Water Pump	2,494
RTI WDP	9,946
RTI DSRP	633
AACRP	31,991
AACRP and NGCC CO ₂ Compressor Auxiliaries	64,795
Recycle Gas Compressor	2,994
Air Cooler Fans	1,657
Water Treatment	2,749
Miscellaneous BOP	5,000
Circulating Water Pumps	7,232
Ground Water Pumps	788
Cooling Tower Fans	4,715
SUBTOTAL CTM PLANT AUXILIARIES, kWe	312,504
NGCC Plant Auxiliary Load Summary, kWe	
Gas Turbine Auxiliaries	787
Steam Turbine Auxiliaries	87
Transformer Losses	2,014
Miscellaneous BOP	500
Air Cooled Condenser Fans	4,273
Condensate Pumps	651
Boiler Feed Water Pumps	15,169
Amine System Auxiliaries	4,112
SCR	16
SUBTOTAL POWER PLANT AUXILIARIES, kWe	27,608
TOTAL AUXILIARIES, kWe	340,113
NET PLANT POWER, kWe	26,948

Table 4-10 shows the carbon balance for the Case 3b plant. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal and natural gas feedstock. Carbon in the air is not part of the carbon capture equation, but is not neglected in the balance since the model

accounts for the air components throughout. Carbon leaves the CTM plant as unburned carbon (minimal for Case 3b) in the slag, CO₂ in the dryer exhaust gas, ASU vent gas, crude methanol product, and the AACRP CO₂ capture product. Additionally, carbon also leaves the NGCC as CO₂ in the stack and Econamine CO₂ capture product. Carbon in the crude methanol is considered as product, not emissions. The carbon capture efficiency is defined as the amount of carbon in the product streams, which include the crude methanol, the dried and compressed CO₂ products from the AACRP and Econamine CO₂ capture processes and gypsum, relative to the amount of carbon in the coal and natural gas feedstock less carbon contained in the slag. For Case 3b, the carbon capture efficiency is 93.4%. **This is greater than the Case 3a carbon capture efficiency, at 92.1%.**

Table 4-10
Case 3b Plant – Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	765,169	
ASU Air	506	
Coal Dryer Air	72	
Power Cycle Natural Gas	46,182	
Power Cycle Combustion Air	450	
DSRP Air In	18	
ASU Vent		506
Coal Dryer Exhaust Gas		46,140
Carbon in Slag		0
Sulfur Product		0
DSRP Vent		2,636
Gypsum		1,205
Crude Methanol Product		342,579
AACRP CO ₂ Product		366,767
NGCC CO ₂ Product		47,184
NGCC Exhaust Gas		5,243
CO ₂ in LTGC Condensate (Vented)		151
Convergence Tolerance		-13
Total	812,398	812,398

Table 4-11 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the elemental sulfur recovered in the DSRP, sulfur recovered in the form of gypsum leaving the lime scrubber, sulfur emitted in the coal dryer exhaust gas, and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible. The net sulfur emissions include only the sulfur emitted in the coal dryer exhaust gas. Based on this, the net sulfur emissions for the plant are < 0.0001 lb of SO₂/MMBtu. **This level of sulfur emissions not only meets the BACT environmental design basis in Table 2-7, but is over 200 times lower than the sulfur emissions for Case 3a. This demonstrates that RTI's advanced warm syngas cleanup process results in a significant decrease in sulfur emissions. At this low level, no additional sulfur removal would be expected for permitting this plant.**

Table 4-11
Case 3a Plant – Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	11,118	
Sulfur Product		4,167
Gypsum		6,928
Coal Dryer Exhaust Gas		0
AACRP CO ₂ Product		21
Convergence Tolerance		1
Total	11,118	11,118

Table 4-12 shows the overall water balance for the Case 3b plant. Raw water withdrawal is the amount of raw water consumed by the plant. The raw water is obtained from groundwater (50%) and from municipal sources (50%). Some water is discharged by the processes as effluent suitable for internal consumption. This effluent is internally recycled and consumed by the slag handling process, as wash water to the AACRP, and as makeup to the cooling tower. Some water is discharged from the process to a permitted outfall. Waste water is discharged by the slag handling process and as cooling tower blowdown.

Table 4-12
Case 3b Plant – Overall Water Balance

Water Use, gpm	Raw Water Withdrawal	Process Effluent Produced for Internal Consumption	Internal Consumption	Process Water Discharge
Slag Handling	0	0	324	(324)
Quench Cooler	1,939	0	0	0
DSRP Air Compressor Knockout	0	(1)	0	0
Syngas Cooling Knockout	0	(2,410)	0	0
AACRP Wash Water	0	0	128	0
CO ₂ Compression Knockout	0	(118)	0	0
Steam Cycle Makeup	1,637	0	0	0
Steam Cycle Blowdown	0	(29)	0	0
Flue Gas Cooling Knockout	0	(193)	0	0
Cooling Tower Makeup	5,074	0	2,300	0
Cooling Tower Blowdown	0	0	0	(1,671)
Total	8,650	(2,752)	2,752	(1,995)

Positive values represent consumption while negative values represent production

4.5 EQUIPMENT LIST

The equipment list for the AR/RTI processes that differ from the reference Shell with Rectisol[®] AGR CTM plant are shown in Table 4-13 below.

Table 4-13
Case 3b Equipment List

ACCOUNT 2 COAL PREPARATION AND FEED
Subaccount 2.3 Dry Coal Injection System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	AR DSP Feed System	AR Proprietary dry solids pump with coal feed bins	1,000 tpd	15	0

ACCOUNT 4 GASIFIER, ASU AND ACCESSORIES
Subaccount 4.1 Gasifier & Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	Gasifier	Vertical, down-fired gasifier with multiple fuel injection ports and including water quench	3,000 tpd	5	0
2	Candle Filters	Ceramic	10,500 acfm	5	0
3	Syngas Cyclone	High efficiency	600,000 lb/hr	5	0

ACCOUNT 5A GAS CLEANUP & PIPING
Subaccount 5A.1 AACRP CO₂ Removal System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	Rich Amine Absorber	Packed Column	228,500 cuft/hr syngas	4	0
2	Lean Amine Absorber	Packed Column w/ Wash Trays	411,000 cuft/hr syngas	2	0
3	Amine Stripper	Packed Column	965,000 cuft/hr gas flow	2	0
4	LP Flash	Packed Column	2,788,700 cuft/hr gas flow	4	0
5	LP Flash Reflux Drum	Reflux Drum	2,753,000 cuft/hr CO ₂	4	0
6	Lean/Rich Amine Exchanger	Shell and Tube	80.7 MMBtu/hr	2	0
7	Amine Stripper Reboiler	Kettle Reboiler	91.9 MMBtu/hr	2	0
8	Lean Amine Cooler	Shell and Tube	46.6 MMBtu/hr	2	0
9	Flash Overhead Condenser	Shell and Tube	20.0 MMBtu/hr	4	0
10	Lean Amine Pump	Centrifugal	1,380 gpm	4	0
11	Semi-lean Amine Pump	Centrifugal	12,000 gpm	4	1

12	Rich Amine Pump	Centrifugal	2,700 gpm	2	0
13	Flash Reflux Pump	Centrifugal	34 gpm	4	0
14	Amine Stripper Bottoms Pump	Centrifugal	2,620 gpm	2	1
15	Filter Pump	Centrifugal	290 gpm	2	0
16	MP Flash Drum	Reflux Drum	111,200 cuft/hr	2	0
17	Recycle Gas Compressor	Centrifugal	122 acfm @ 456 psia	2	0
18	Amine Tank	Vertical Vessel	7,500 gallons	2	0
19	Filter Package	Filter Package	291 gpm	2	0
18	Filter Package Vessel	Vessel	1,500 gallons	3	0

ACCOUNT 5A GAS CLEANUP & PIPING
Subaccount 5A.2 RTI DSRP System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	DSRP Fixed Bed Reactor	Packed Bed Reactor	338,000 lb/hr	1	0
2	Tail Gas Oxidation Reactor	Packed Bed Reactor	380,100 lb/hr	1	0
3	DSRP Feed Gas Preheater	Shell and Tube	17 MMBtu/hr	1	0
4	Sulfur Condenser	Shell and Tube	4,100 lb/hr sulfur	1	0
5	Oxidation Reactor Gas Preheater	Shell and Tube	7.9 MMBtu/hr	1	0
6	Liquid Sulfur Separator	Pressure Vessel	4,100 lb/hr sulfur	1	0
7	Recycle Gas Compressor	Centrifugal	1,200 acfm @ 750 psia	1	0
8	Pulsing Gas Compressor	Centrifugal	Average 15 acfm @ 815 psia	1	0
9	Scrubber Recycle Cooler	Shell and Tube	76 MMBtu/hr	1	0
10	SO ₂ Scrubber	Tray Column	396,000 lb/hr	1	0
11	Recycle Pump	Centrifugal	1,915 gpm @ 288 ft H ₂ O	1	0
12	Lime Slurry Pump	Centrifugal	350 gpm @ 1,617 ft H ₂ O	1	0
13	Gypsum Filter	Bag Filter	1,000 gpm	2	0
14	Lime Makeup Drum	Pressure Vessel	175,000 lb/hr	1	0

ACCOUNT 5A GAS CLEANUP & PIPING
Subaccount 5A.4a RTI WDP System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	WDP Adsorption Reactor	Proprietary Transport Reactor	38,340 acfm	1	0

2	WDP Regeneration Reactor	Proprietary Transport Reactor	1,140 acfm	1	0
3	Adsorber Cyclone Separator	Cyclone Separator	38,340	1	0
4	Stripper Cyclone Separator	Cyclone Separator	1,140 acfm	1	0
5	Adsorber Filter	Candle Filter	38,340 acfm	1	0
6	Regenerator Offgas Filter	Candle Filter	1,140 acfm	1	0
7	Adsorber Filter Lock Hopper + Fines Bin		Proprietary	1	0
8	Regenerator Filter Lock Hopper + Fines Bin		Proprietary	1	0
9	Sorbent Feeder Package (incl Hopper)		Proprietary	1	0
10	Regenerator Air Heat Exchanger	Shell and Tube	73,300 lb/hr	1	0
11	Recycle Syngas Compressor	Centrifugal	240 acfm @ 829 psia	1	0
12	Regenerator Air Compressor	Centrifugal	30,000 acfm @ 15 psia	1	0
13	DSRP Offgas Compressor	Centrifugal	160 acfm @ 810 psia	1	0
14	Syngas Recycle Compressor	Centrifugal	168 acfm @ 898 psia	1	0

ACCOUNT 5A SYNGAS CLEANUP
Subaccount 5A.4b Shift Reactors

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	AFWGS Reactor 1	Packed Bed Reactor	Proprietary	2	0
2	AFWGS Reactor 2	Packed Bed Reactor	Proprietary	2	0

4.6 CAPITAL COST

Table 4-14 shows the cost breakdown of the Case 3b plant, consistent with the code of accounts format as expressed in the Crude Methanol Study. The accounts/subaccounts of interest for this study are:

- 2.3 Dry Coal Injection System,
- 4 Gasifier & Accessories,
- 5A Gas Cleanup & Piping and
- 5B CO₂ Removal and Compression

These are shown with more detail to include the various subaccounts and provide more clarity to the major cost differences among the cases.

Table 4-15 shows the calculation and addition of owner's costs to determine the TOC, which is used to calculate the product methanol RSP.

The estimated TOC of the Case 3b plant in 2011 dollars is \$4,567MM.

Table 4-14
Case 3b Plant – Total Plant Cost Summary

Case 3b: AR Gasifier with RTI WDP CTM Plant			Coal Feed, lb/hr		1,528,260		Plant Size		10079 metric tons per day			
Low Rank Western Coal Baseline Study			PRB		Coal HHV, Btu/lb		8,564					
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost \$	Eng'g CM H.O & Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect				Process	Project	\$	
1	COAL & SORBENT HANDLING	\$35,251	\$6,321	\$27,325	\$0	\$0	\$68,897	\$6,253	\$0	\$15,030	\$90,180	
2	COAL & SORBENT PREP & FEED											
2.3	Dry Coal Injection System	\$134,811	\$0	\$88,517	\$0	\$0	\$223,328	\$19,235	\$44,666	\$57,446	\$344,674	
2.x	Other Coal & Sorbent Prep & Feed Systems	\$146,191	\$27,729	\$42,212	\$0	\$0	\$216,131	\$18,906	\$0	\$47,008	\$282,045	
	SUBTOTAL 2.	\$281,001	\$27,729	\$130,729	\$0	\$0	\$439,459	\$38,141	\$44,666	\$104,454	\$626,719	
3	FEEDWATER & MISC BOP SYSTEMS	\$28,134	\$7,918	\$16,967	\$0	\$0	\$53,019	\$4,981	\$0	\$13,176	\$71,176	
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries (AR)	\$88,269	\$44,134	\$133,274	\$0	\$0	\$265,677	\$23,821	\$66,419	\$53,388	\$409,304	
4.3	ASU/Oxidant Compression	\$539,576	\$0	\$0	\$0	\$0	\$539,576	\$52,301	\$0	\$59,187	\$651,064	
4.4	LT Heat Recovery & FG Saturation	\$76,936	\$0	\$24,424	\$0	\$0	\$101,360	\$9,892	\$0	\$22,250	\$133,502	
4.6	Flare Stack System	\$0	\$1,421	\$647	\$0	\$0	\$2,068	\$198	\$0	\$453	\$2,719	
4.9	Gasification Foundations	\$0	\$30,489	\$18,183	\$0	\$0	\$48,671	\$4,456	\$0	\$13,282	\$66,409	
	SUBTOTAL 4.	\$704,781	\$76,044	\$176,528	\$0	\$0	\$957,352	\$90,668	\$66,419	\$148,560	\$1,262,999	
5A	GAS CLEANUP & PIPING											
5A.1	AACRP	\$38,471	\$0	\$32,316	\$0	\$0	\$70,787	\$6,688	\$14,157	\$18,326	\$109,959	
5A.2	RTI DSRP System	\$16,676	\$0	\$18,461	\$0	\$0	\$35,137	\$3,413	\$7,027	\$9,115	\$54,692	
5A.3	Mercury Removal	\$7,303	\$0	\$5,520	\$0	\$0	\$12,822	\$1,238	\$641	\$2,940	\$17,641	
5A.4a	RTI WDP System	\$45,217	\$0	\$54,260	\$0	\$0	\$99,476	\$9,672	\$19,895	\$25,809	\$154,853	
5A.4b	Shift Reactors	\$8,455	\$0	\$3,382	\$0	\$0	\$11,837	\$1,135	\$0	\$2,595	\$15,567	
5A.5	Syngas Compressor	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
5A.6	Blowback Gas Systems	\$6,453	\$1,086	\$612	\$0	\$0	\$8,150	\$773	\$0	\$1,785	\$10,708	
5A.7	Fuel Gas Piping	\$0	\$2,884	\$1,887	\$0	\$0	\$4,771	\$442	\$0	\$1,042	\$6,255	
5A.9	HGCU Foundations	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
	SUBTOTAL 5.	\$122,575	\$3,970	\$116,437	\$0	\$0	\$242,981	\$23,361	\$41,721	\$61,613	\$369,676	
5B	CO2 REMOVAL & COMPRESSION											
5B.1	NGCC CO2 Removal System	\$23,700	\$0	\$7,142	\$0	\$0	\$30,843	\$2,573	\$6,168	\$7,917	\$47,501	
5B.2	CO2 Compression & Drying (AACRP & NGCC)	\$68,653	\$0	\$28,881	\$0	\$0	\$94,534	\$9,057	\$0	\$20,718	\$124,309	
5B.3	CO2 Compression & Drying (NGCC)	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
	SUBTOTAL 5B.	\$92,353	\$0	\$36,023	\$0	\$0	\$128,376	\$11,630	\$6,168	\$28,635	\$171,810	
5C	METHANOL PRODUCTION	\$131,611	\$55,371	\$110,743	\$0	\$0	\$297,725	\$29,773	\$0	\$65,500	\$392,998	
6	COMBUSTION TURBINE/ACCESSORIES	\$47,397	\$394	\$3,814	\$0	\$0	\$51,605	\$11,171	\$11,909	\$14,852	\$89,537	
7	HRSG, DUCTING & STACK	\$21,496	\$601	\$5,106	\$0	\$0	\$27,203	\$2,259	\$0	\$2,962	\$32,424	
8	STEAM TURBINE GENERATOR	\$78,650	\$1,021	\$16,025	\$0	\$0	\$95,697	\$8,692	\$0	\$18,869	\$123,258	
9	COOLING WATER SYSTEM	\$18,105	\$19,849	\$14,701	\$0	\$0	\$52,655	\$4,871	\$0	\$11,695	\$69,221	
10	ASH/SPENT SORBENT HANDLING SYS	\$77,771	\$3,478	\$73,220	\$0	\$0	\$154,469	\$14,945	\$0	\$17,687	\$187,101	
11	ACCESSORY ELECTRIC PLANT	\$37,308	\$21,790	\$38,364	\$0	\$0	\$97,462	\$8,474	\$0	\$21,084	\$127,020	
12	INSTRUMENTATION & CONTROL	\$12,333	\$3,728	\$10,344	\$0	\$0	\$26,404	\$2,371	\$1,320	\$5,393	\$35,488	
13	IMPROVEMENTS TO SITE	\$5,761	\$2,831	\$15,078	\$0	\$0	\$23,670	\$2,337	\$0	\$7,802	\$33,809	
14	BUILDINGS & STRUCTURES	\$0	\$7,648	\$10,601	\$0	\$0	\$18,250	\$1,649	\$0	\$3,230	\$23,129	
CALCULATED TOTAL COST		\$1,694,528	\$238,692	\$802,004	\$0	\$0	\$2,735,224	\$261,575	\$172,203	\$540,541	\$3,706,544	

Table 4-15
Case 3b Plant – Total Overnight Cost Breakdown

Owner's Costs	\$ x \$1,000
Preproduction Costs	
6 months All Labor	\$26,844
1 Month Maintenance Materials	\$4,961
1 Month Non-Fuel Consumables	\$1,823
1 Month Waste Disposal	\$1,154
25% of 1 Months Fuel Cost at 100% CF	\$6,716
2% of TPC	\$74,131
Total	\$115,628
Inventory Capital	
60 day supply of fuel at 100% CF	\$52,993
60 day supply of non-fuel consumables at 100% CF	\$2,262
0.5% of TPC (spare parts)	\$18,533
Total	\$73,788
Initial Cost for Catalyst and Chemicals	\$13,978
Land	\$900
Other Owner's Cost	\$555,982
Financing Costs	\$100,077
Total Owner's Costs	\$860,352
Total Overnight Costs (TOC)	\$4,566,896

4.7 OPERATING COSTS

Table 4-16 shows the OPEX breakdown for the Case 3b plant.

Table 4-16
Case 3b Plant – Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Case 3b - AR Gasifier with RTI WDP CTM Plant				
Plant Size (mtpd Methanol)	10,079				
Primary/Secondary Fuel:	PRB/Natural Gas				
Design/Construction	5 years		Book Life (yrs):	20	
TPC (Plant Cost) Year	June 2011		TPI Year:	2016	
Capacity Factor (%)	90		CO2 Captured (TPD)	16121	
OPERATING & MAINTENANCE LABOR					
Operating Labor					
Operating Labor Rate (base):	\$39.70 \$/hr				
Operating Labor Burden:	30.00 % of base				
Labor Overhead Charge	25.00 % of labor				
Operating Labor Requirements per Shift	units/mod		Total Plant		
Skilled Operator	2.0		2.0		
Operator	10.0		10.0		
Foreman	1.0		1.0		
Lab Tech's etc	3.0		3.0		
TOTAL Operating Jobs	16.0		16.0		
Annual Cost					
\$					
Annual Operating Labor Cost	\$7,233,658				
Maintenance Labor Cost	\$35,716,254				
Administration & Support Labor	\$10,737,478				
Property Taxes and Insurance	\$74,130,871				
TOTAL FIXED OPERATING COSTS	\$127,818,260				
VARIABLE OPERATING COSTS					
Maintenance Material Cost					
					\$53,574,380
Consumables	Consumption	Unit	Initial Fill		
	Initial	/Day	Cost	Cost	
Water(/1000 gallons)	0	6,228	1.67	\$0	\$3,424,561
Chemicals					
MU & WT Chem (lb)	0	37107	0.27	\$0	\$3,264,984
Carbon (Hg Removal) (lb)	249237	380	1.63	\$406,256	\$203,290
MEA Solvent (ton)	160	0.22	3751.70	\$601,679	\$270,293
Water Gas Shift Catalyst (ft3)	6606	3.62	394.60	\$2,606,896	\$469,241
WDP Sorbent (lb)	Proprietary	Proprietary	Proprietary	\$6,601,715	\$5,859,411
AACRP Solvent	327949	179.70	2.80	\$917,558	\$165,160
Methanol Synthesis Catalyst	4011	3.66	534.68	\$2,144,768	\$642,717
DSRP Catalyst	54120	29.65	11.15	\$603,438	\$108,619
Oxidation/Reduction Catalyst	60940	33.39	1.56	\$95,066	\$17,112
Calcium Hydroxide	5200	419040	0.04	\$199	\$5,259,942
Subtotal Chemicals				\$13,977,576	\$16,260,769
Other					
Supplemental Electricity (MWh consumed)	0	0	62.33	\$0	\$0
Gases, N2 etc./100scf	0	0	0.00	\$0	\$0
LP Steam (/1000 lbs)	0	0	0.00	\$0	\$0
Subtotal Other				\$0	\$0
Waste Disposal:					
Spent Mercury Catalyst (lb)	0	380	0.65	\$0	\$81,067
Flyash (ton)	0	0	0.00	\$0	\$0
Slag (ton)	0	1502	25.11	\$0	\$12,385,561
Subtotal Waste Disposal				\$0	\$12,466,628
By-products & Emissions					
Sulfur (tons)	0	0.0	0.00	\$0	\$0
Supplemental Electricity (MWh generated)	0	647	-59.59	\$0	-\$12,659,876
Subtotal By-Products				\$0	-\$12,659,876
TOTAL VARIABLE OPERATING COSTS				\$13,977,576	\$73,066,462
Coal (tons)	0	18,339	36.57	\$0	\$220,312,342
Natural Gas (MMBtu)	0	34,675	6.13	\$0	\$69,825,571

4.8 METHANOL PRODUCT REQUIRED SELLING PRICE

Table 4-17 shows a summary of the power output, CAPEX, OPEX, and methanol product RSP for the Case 3b plant. The Case 3b plant methanol RSP is estimated to be \$1.21/gal under the loan guarantee finance structure and \$1.40/gal under the commercial fuels finance structure.

Table 4-17
Case 3b Plant – Overall Performance and Economic Summary

Case	Case 3b	
CAPEX, \$MM		
Total Installed Cost (TIC)	\$2,732	
Total Plant Cost (TPC)	\$3,707	
Total Overnight Cost (TOC)	\$4,567	
OPEX, \$MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC _{fix})	\$128.0	
Variable Operating Cost, less Fuel (OC _{var})	\$95.3	
Coal Feedstock (OC _{coal})	\$244.8	
Natural Gas Feedstock (OC _{NG})	\$77.6	
Import/(Export) Power (OC _{Power})	(\$14.1)	
Total OPEX	\$531.4	
Plant Output		
Crude Methanol Product, tons per year	4,055,274	
Net Power Output, MWe	26.95	
Required Selling Price^A		
Excluding CO₂ TS&M^{BE}, \$/ton	347.3	407.3
Including CO₂ TS&M^{BE}, \$/ton	363.6	423.7
RSP Component Details (\$/gal)		
Capital ^B	0.70	0.90
Fixed O&M	0.12	
Variable O&M	0.08	
Coal	0.20	
Natural Gas	0.06	
Power	(0.01)	
CO ₂ TS&M	0.05	
RSP^B Total (\$/gal)	1.21	1.40
Costs of CO₂ Captured^{B,C} (\$/tonne)	18.9	21.1
Costs of CO₂ Avoided^{B,D} (\$/tonne)	29.9	32.1

^A Capacity factor assumed to be 90 percent

^B Values shown are for two financial structures

The first (lower value) is based on the loan guarantee finance structure

The second (higher value) is based on the commercial fuels finance structure

^C Excludes CO₂ TS&M

^D Includes CO₂ TS&M

^E Based on 301.73 gallons/ton

Section 5 Case 3c: AR Compact Gasifier with Rectisol® AGR CTM Plant

5.1 PROCESS OVERVIEW

The Case 3c AR Compact Gasifier with Rectisol® AGR CTM plant, like the previous two cases, produces a nominal 10,000 metric tons per day of methanol by converting Montana PRB coal to syngas of appropriate quality via gasification, WGS and AGR. The syngas is then converted to methanol by catalytic conversion in methanol synthesis reactors. This Case 3c plant is equipped with an NGCC to generate enough power to meet the plant's auxiliary demands such that no power needs to be imported, and any excess power is exported for extra revenue. The Case 3c plant is designed to capture CO₂ with a carbon capture efficiency of more than 90% of the carbon in the coal and natural gas. The Case 3c plant operates with an annual on-stream CF of 90%.

Case 3c (AR gasifier) and Case 3d (RTI advanced warm gas cleanup) enable evaluation of the performance and cost for the individual advanced technologies (AR gasifier and RTI warm syngas cleanup) with conventional technologies for the balance of the CTM plant. These cases can then be compared with Case 3b to evaluate if there are any synergistic advantages in integrating both systems in a CTM plant.

The Case 3c plant has the following characteristics that differentiate it from the Case 3a Reference plant:

- The AR DSP feed system replaces the reference case's lockhopper system used to feed dried coal to the gasifier. Nexant had previously evaluated the AR DSP feed system in comparison to the Shell lockhopper feed system in a separate study (DE-FE0012062) for DOE on behalf of AR. The results from the earlier study are used to establish the performance and cost of the DSP feed system for this case.
- The AR compact gasifier replaces the Shell gasifier from Case 3a.
- The AR DSP is able to achieve a higher outlet pressure for the coal particles leaving the system. Therefore, the operating pressures for all systems upstream of the methanol synthesis reactors are raised. This eliminates the syngas booster compressor between the Rectisol® unit and the methanol synthesis reactors and its associated power consumption.

The coal feed rate is the same as Case 3b since both utilize the AR gasification process operating under the same conditions, hence same cold gas efficiency, to generate enough syngas for production of 10,000 mtpd of crude methanol.

5.2 CASE 3C PROCESS DESCRIPTION

The overall BFD for the Case 3c plant is shown in Figure 5-1, with the accompanying stream flows shown in Table 5-1. Additional descriptions of the Case 3c plant's various processes are provided below. To better visualize the different unit operations in the CTM plant, simplified PFDs of the various plant processes are depicted in Figures 5-2 through 5-9. Tables 5-2 through 5-9 provide the model-generated process data for the numbered streams referenced in the PFDs.

Coal Preparation and Drying

The description is the same as Case 3a. However, due to the larger fuel gas heating value supplied to the coal dryer, no supplemental natural gas firing is required. Figure 5-2 depicts the coal drying process while Table 5-2 presents the mass balances.

AR Dry Solids Pump (DSP) Coal Feed System

Same as Case 3b.

Air Separation Unit

Same description as Case 3a, but the oxygen for the gasifier is supplied at 1,000 psia rather than 711 psia.

AR Compact Gasifier

Same description as Case 3b. The coal feed, ASU, gasification and quench processes are shown in Figure 5-3 and their mass balances are presented in Table 5-3.

Dry Solids Removal and Wet Scrubbing

Same description as Case 3a, but operating at a higher pressure

Water Gas Shift

Same description as Case 3a but operating at a higher pressure. The wet scrubbing and water-gas shift processes are shown in Figure 5-3 and the corresponding mass balances are presented in Table 5-4.

Like the reference case in the Crude Methanol Study, the Case 3c syngas leaving the first WGS reactor exceeds 900°F. Additionally, the water to dry gas ratio leaving the second reactor is below 0.25. It should be noted that these conditions potentially lead to side reactions that may deactivate the catalyst prematurely and change the product slate. Though not shown in this case, it may be necessary to inject additional steam, which will function as a heat carrier, to decrease the syngas temperature at the first WGS reactor outlet, as well as raise the water to dry gas ratio at the second WGS reactor outlet.

Low Temperature Gas Cooling and Mercury Removal

Same description as Case 3a, but operating at higher pressure. The low temperature gas cooling and mercury removal processes are shown in Figure 5-4 and the corresponding mass balances are presented in Table 5-5.

Acid Gas Removal

The description of the Rectisol[®]-based AGR system is the same as Case 3a with the exception that it is operating at higher pressure. The PFD for the Rectisol[®] and CO₂ compression processes are shown in Table 5-6 and the corresponding mass balances are presented in Table 5-6.

When operating at higher pressures, the Rectisol[®] unit's cost is expected to be reduced, mainly due to the higher CO₂ partial pressure, which aids absorption. The approximate percentage cost decrease for the Rectisol[®] unit operating at 815 psia against the Case 3a unit at 515 psia operation was estimated. This cost decrease for high pressure operation is reflected in Nexant's Case 3c plant Rectisol[®] unit's cost estimate.

CO₂ Compression and Dehydration

Same description as Case 3a. However, as the AR gasifier operates at a higher pressure, CO₂ that is used for conveying the dried coal to the gasifier has to be withdrawn at a higher pressure (1,100 psia) from the CO₂ compression unit.

Sulfur Recovery Unit

Same description as Case 3a. The SRU simplified PFD is shown in Figure 5-7, while the material balance is shown in Table 5-7. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table.

Methanol Reactor and Synthesis Loop

Same description as Case 3b. Note that the syngas feed compressor to the methanol synthesis reactors has been eliminated due to the higher system pressure upstream. Also, there is excess purge gas after taking into account the fuel requirements for coal drying. This excess gas is routed to the NGCC to produce power, thus reducing the natural gas firing rate for power generation. The Case 3c methanol synthesis PFD is shown in Figure 5-8 and the corresponding mass balances are presented in Table 5-8.

In its correspondence with Nexant, Linde specified that a certain amount of makeup methanol was required for continued operation of the Rectisol[®] AGR process. To account for this, the makeup methanol quantity to the Rectisol[®] unit was debited against the gross crude methanol production. Net methanol production from the CTM plant is the gross methanol production rate less makeup Rectisol[®] quantity.

NGCC

Same as Case 3a. The Case 3c NGCC PFD is shown in Figure 5-9 and the corresponding mass balances are presented in Table 5-9.

NGCC Flue Gas Post Combustion CO₂ Capture and Compression

Same as Case 3a

Figure 5-1
Case 3c Plant – Overall BFD

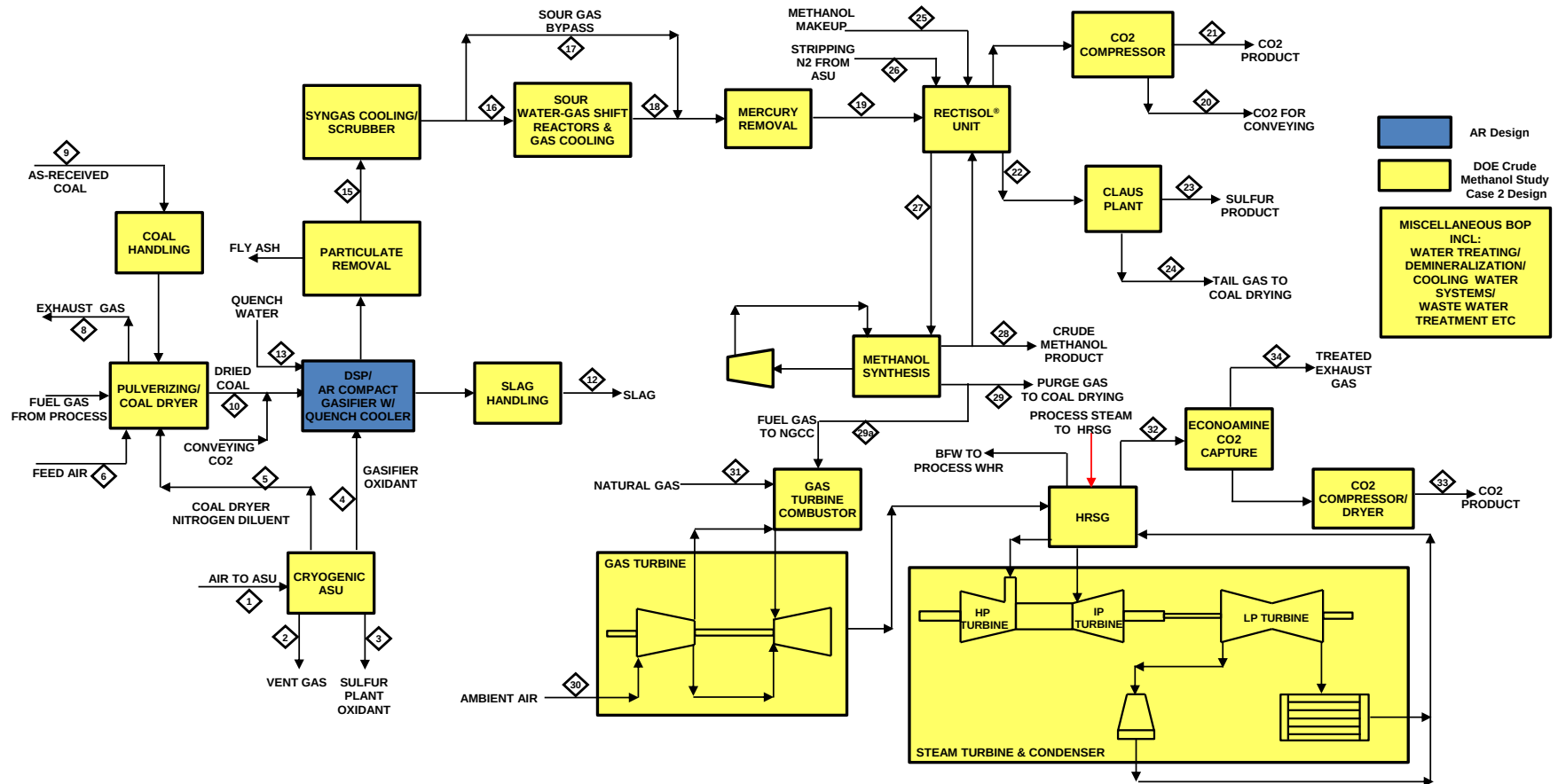


Table 5-1
Case 3c Plant – Overall Stream Table

Description	Air to ASU	ASU Vent Gas	Sulfur Plant Oxidant	Gasifier Oxidant	Nitrogen Diluent to Dryer	Air Feed to Coal Dryer	Coal Dryer Exhaust	As-Received Coal	Dried Coal	Quenched Syngas	Slag	Quench Water	Syngas to Scrubber	Syngas to Shift	Bypass Syngas	Shifted Syngas
Stream No.	1	2	3	4	5	6	8	9	10	11	12	13	15	16	17	18
V-L Mole Fraction																
Ar	0.0092	0.0219	0.0300	0.0300	0.0023	0.0092	0.0077	0	0	0.0049	0	0	0.0049	0.0046	0.0046	0.0046
CH4	0	0	0	0	0	0	0	0	0	0.0022	0	0	0.0022	0.0020	0.0020	0.0020
CO	0	0	0	0	0	0	0	0	0	0.3982	0	0	0.3982	0.3717	0.3717	0.0722
CO2	0.0003	0.0052	0	0	0	0.0003	0	0	0	0.0124	0	0	0.0124	0.0117	0.0117	0.3115
COS	0	0	0	0	0	0	0	0	0	0.0002	0	0	0.0002	0.0002	0.0002	0.0000
H2	0	0	0	0	0	0	0	0	0	0.1612	0	0	0.1612	0.1505	0.1505	0.4501
H2O	0.0099	0.1536	0	0	0	0.0099	0.1568	0	0	0.4126	0	1.0000	0.4126	0.4514	0.4514	0.1516
H2S	0	0	0	0	0	0	0	0	0	0.0019	0	0	0.0019	0.0018	0.0018	0.0020
N2	0.7732	0.6994	0.0200	0.0200	0.9921	0.7732	0.7867	0	0	0.0058	0	0	0.0058	0.0054	0.0054	0.0054
NH3	0	0	0	0	0	0	0	0	0	0.0006	0	0	0.0006	0.0006	0.0006	0.0006
O2	0.2074	0.1200	0.9500	0.9500	0.0054	0.2074	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0.0000	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0	0	1.0000	0	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	129693	8213	249	26489	94152	17591	137034	0	0	161262	0	64000	161262	113933	58754	113933
V-L Flowrate, lb/hr	3742555	224206	8014	851820	2641937	507612	3722996	0	0	3215573	0	1152978	3215573	2257561	1164189	2257581
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	1528260	1206840	0	125127	0	0	0	0	0
Temperature, F	59	64	72	292	70	59	180	59	180	797		392	602	435	435	636
Pressure, psia	14.7	16.4	23.2	1000.0	18.0	14.7	14.3	14.7	14.3	915.0	915.0	980.0	899.0	899.0	899.0	861.0

Table 5-1 (cont'd)
Case 3c Plant – Overall Stream Table

Description	Syngas to Rectisol®	Convey'g CO2	CO2 Product	Rectisol® Tail Gas to Claus	Sulfur Product	Tail Gas to Coal Drying	Methanol Makeup	Stripping N2 from ASU	Treated Syngas to MeOH	Crude Methanol Product	MeOH Synthesis Purge Gas	Excess Purge Gas to NGCC	Air to Gas Turbine	Natural Gas Feed	NGCC Flue Gas Exhaust	CO2 Product	Treated NGCC Exhaust
Stream No.	19	20	21	22	23	24	25	26	27	28	29	29a	30	31	32	33	34
V-L Mole Fraction																	
Ar	0.0062	0.0001	0.0001	0	0	0.0060	0	0.0023	0.0083	0.0001	0.1204	0.1204	0.0092	0	0.0099	0	0.0107
CH4	0.0027	0	0	0	0	0	0	0	0.0036	0	0.0495	0.0495	0	0.9310	0	0	0
CO	0.2330	0.0032	0.0032	0.0024	0	0.0162	0	0	0.3140	0	0.1843	0.1843	0	0	0	0	0
CO2	0.2802	0.9769	0.9769	0.6759	0	0.5226	0	0	0.0347	0.0113	0.2209	0.2209	0.0003	0.0100	0.0370	1.0000	0.0040
COS	0.0001	0	0	0	0	0.0008	0	0	0.0000	0	0	0	0	0	0	0	0
H2	0.4661	0.0004	0.0004	0.0020	0	0.0072	0	0	0.6300	0	0.2635	0.2635	0	0	0	0	0
H2O	0.0019	0	0	0	0	0.4060	0	0	0	0.0114	0.0001	0.0001	0.0099	0	0.0758	0	0.0328
H2S	0.0026	0	0	0.3021	0	0.0026	0	0	0	0	0	0	0	0	0	0	0
N2	0.0072	0.0188	0.0188	0.0025	0	0.0349	0	0.9921	0.0093	0.0000	0.1412	0.1412	0.7732	0.0160	0.7449	0	0.8086
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0.0001	0.0001	0	0	0.0003	0	0.0054	0	0	0	0	0.2074	0	0.1325	0	0.1438
SO2	0	0	0	0	0	0.0013	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	1.0000	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0.0011	0	0.0008	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol	0	0.0002	0.0002	0	0	0	1.0000	0	0.0001	0.9755	0.0201	0.0201	0	0	0	0	0
Ethanol	0	0	0	0	0	0	0	0	0	0.0014	0	0	0	0	0	0	0
Propanol	0	0	0	0	0	0	0	0	0	0.0001	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0	0	0	0	0	0	0.032	0	0	0
Propane	0	0	0	0	0	0	0	0	0	0	0	0	0	0.007	0	0	0
n-Butane	0	0	0	0	0	0	0	0	0	0	0	0	0	0.004	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	128996	2959	29907	1090	340	1448	17	591	95402	29258	4683	993	114330	3730	118938	3956	109566
V-L Flowrate, lb/hr	2633987	129095	1304832	44720	10900	46743	551	16584	1168052	937317	119992	25453	3299204	64622	3388952	174088	3117277
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature, F	111	269	162	59	320	320	123	86	81	118	130	130	59	60	298	124	85
Pressure, psia	815.0	1100.0	2215	28	20	20	40	75.0	785	40	717	717	14.5	474.7	14.9	2215	15.0

Figure 5-2
Case 3c Plant – Coal Drying PFD

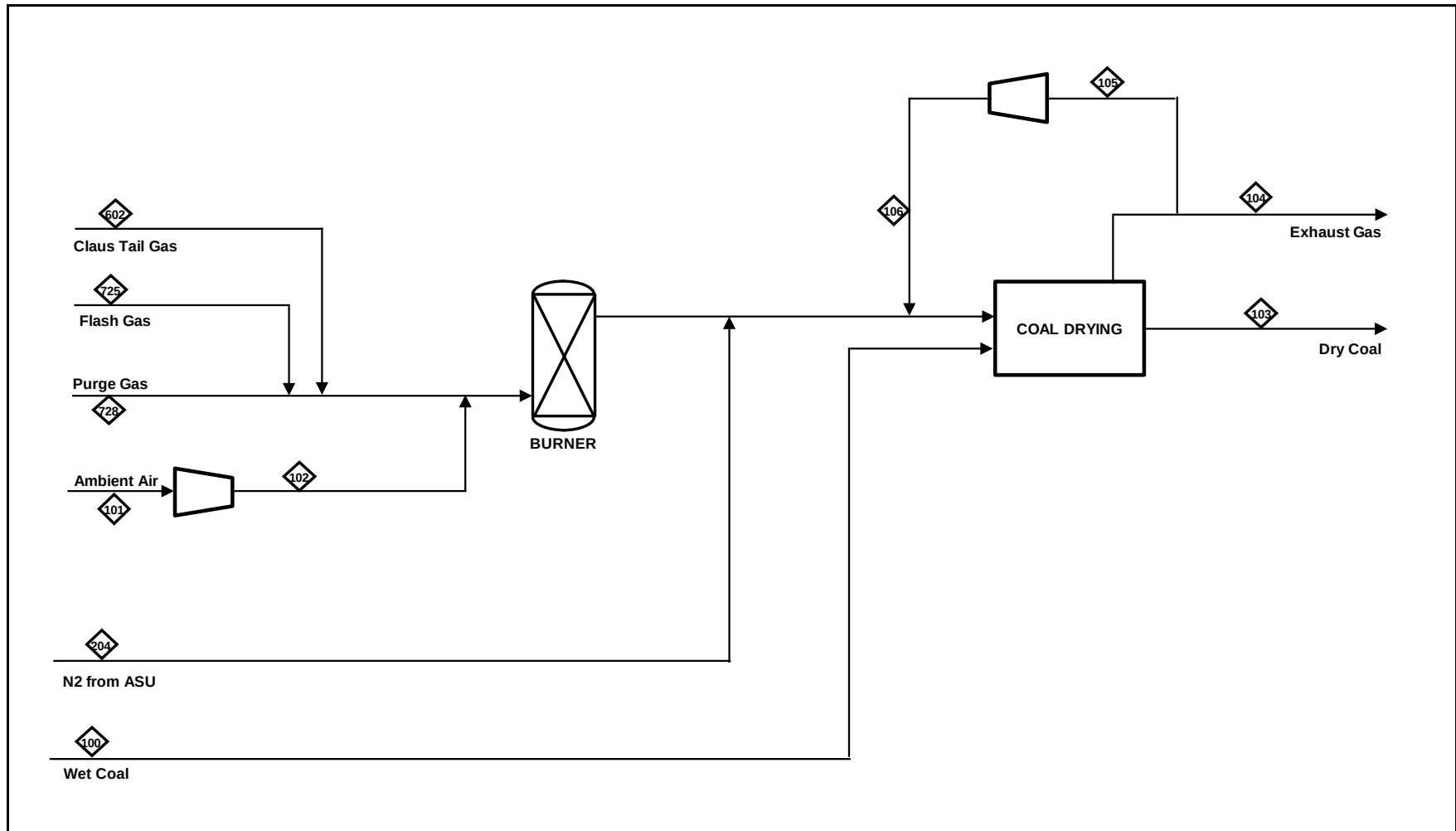


Table 5-2
Case 3c Plant – Coal Drying Stream Table

STREAM	100	101	102	103	104	105	106	204	602	725	728
Description	Wet Coal	Ambient Air	Feed Air	Dry Coal	Exhaust Gas	LP Recycle Gas	HP Recycle Gas	Dilution Nitrogen	Claus Tail Gas	MeOH Flash Gas	MeOH Purge Gas
Mole Flow (Vapor/Liquid)/lbmol/hr											
AR	0	163	163	0	1056	569	569	217	9	105	564
CH4	0	0	0	0	0	0	0	0	1	62	232
CO	0	0	0	0	0	0	0	0	23	55	863
CO2	0	6	6	0	4968	2675	2675	0	757	1432	1035
COS	0	0	0	0	0	0	0	0	1	0	0
H2	0	0	0	0	0	0	0	0	10	33	1234
H2O	0	174	174	0	21494	11574	11574	19	588	2	0
H2S	0	0	0	0	0	0	0	0	4	0	0
N2	0	13600	13600	0	107809.3	58051	58051	93408.11	51	89	661
NH3	0	0	0	0	0	0	0	0	0	0	0
O2	0	3648	3648	0	1700	915	915	508	0	0	0
SO2	0	0	0	0	7	4	4	0	2	0	0
S	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	1	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0	2	404	94
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0.4	0.1
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	0	17591	17591	0	137034	73788	73788	94152	1448	2181	4683
Total V/L Flow, lb/hr	0	507612	507612	0	3722996	2004693	2004693	2641937	46743	85274	119992
Solids Mass Flow, lb/hr	1528260	0	0	1206840	0	0	0	0	0	0	0
Coal	1528260	0	0	1206840	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	1528260	507612	507612	1206840	3722996	2004693	2004693	2641937	46743	85274	119992
Pressure, psia	14.7	14.7	17	14.3	14.3	14.3	16	18	20	40	717
Temperature, F		59	90		180	180	208	70	320	118	130

Figure 5-3
Case 3c Plant – ASU/Coal Gasification PFD

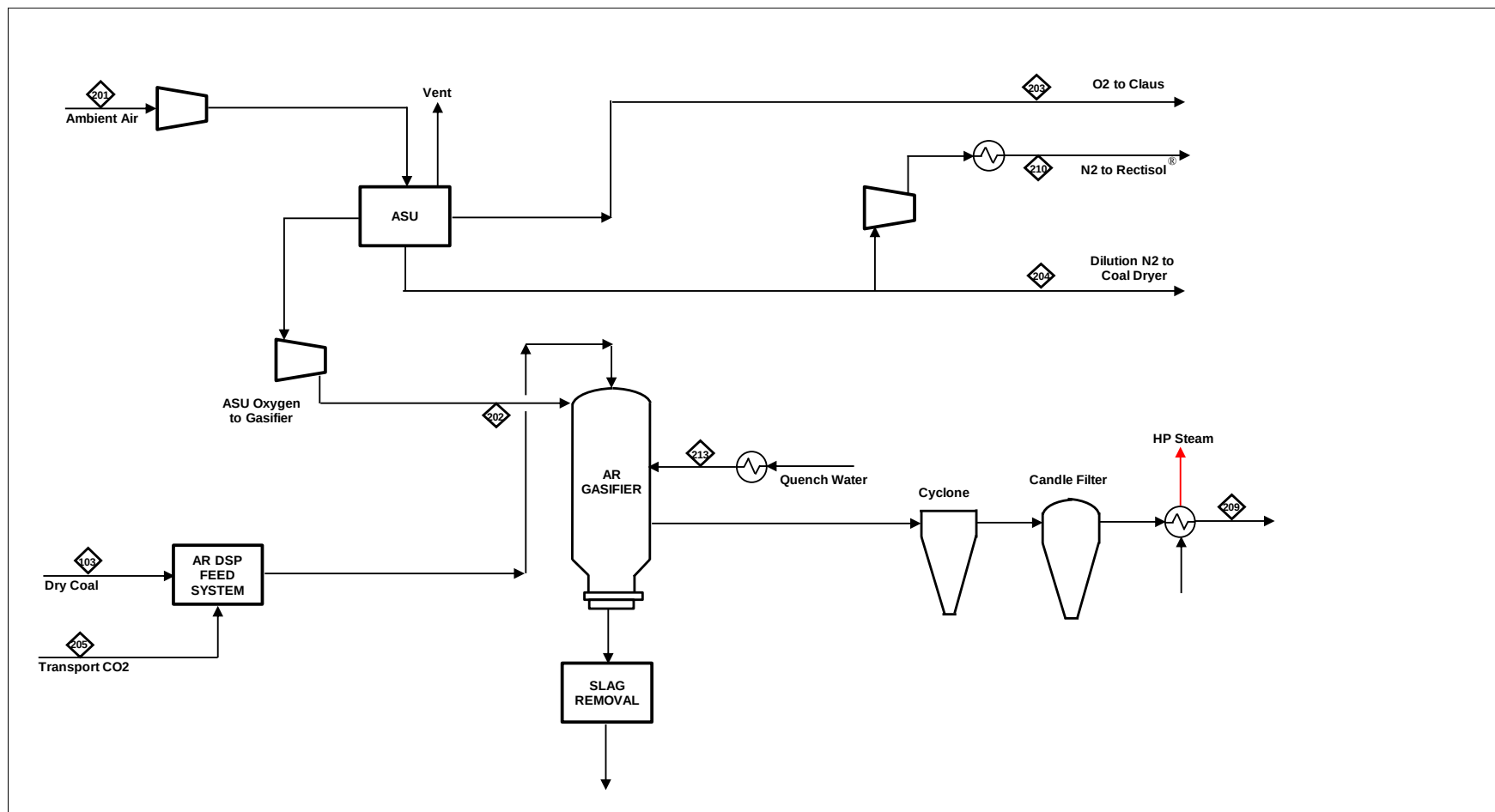


Table 5-3
Case 3c Plant – ASU/Coal Gasification Stream Table

STREAM	201	202	203	204	205	103	318	213	214	209	210
Description	Ambient Air	Oxygen to Gasifier	Oxygen to Claus	N2 to Coal Dryer	Conveying CO2	Dry Coal	Recycle Syngas	Quench Water	Slag Out	Raw Cooled Syngas	Stripping N2 to Rectisol [®]
Mole Flow (Vapor/Liquid)/lbmol/hr											
AR	1199	794	7	217	0	0	0	0	0	795	1
CH4	0	0	0	0	1	0	0	0	0	353	0
CO	0	0	0	0	9	0	0	0	0	64212	0
CO2	43	0	0	0	2890	0	0	0	0	2002	0
COS	0	0	0	0	0	0	0	0	0	36	0
H2	0	0	0	0	1	0	0	0	0	25998	0
H2O	1280	0	0	19	0	0	0	64000	0	66531	0
H2S	0	0	0	0	0	0	0	0	0	311	0
N2	100273	530	5	93408	56	0	0	0	0	928	586
NH3	0	0	0	0	0	0	0	0	0	90	0
O2	26898	25165	237	508	0	0	0	0	0	0	3
SO2	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	3	0
HCL	0	0	0	0	0	0	0	0	0	3	0
Methanol -- CH3OH	0	0	0	0	1	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	129693	26489	249	94152	2959	0	0	64000	0	161262	591
Total V/L Flow, lb/hr	3742555	851820	8014	2641937	129095	0	0	1152978	0	3215573	16584
Solids Mass Flow, lb/hr											
Coal	0	0	0	0	0	1206840	0	0	125127	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	3742555	851820	8014	2641937	129095	1206840	0	1152978	125127	3215573	16584
Pressure, psia	14.7	1000	23.2	18.0	1100	14.3	0	980	915	899	75
Temperature, F	59	292	72	70	269		0	392		602	86

Figure 5-4
Case 3c Plant – Wet Scrubber/Water Gas Shift PFD

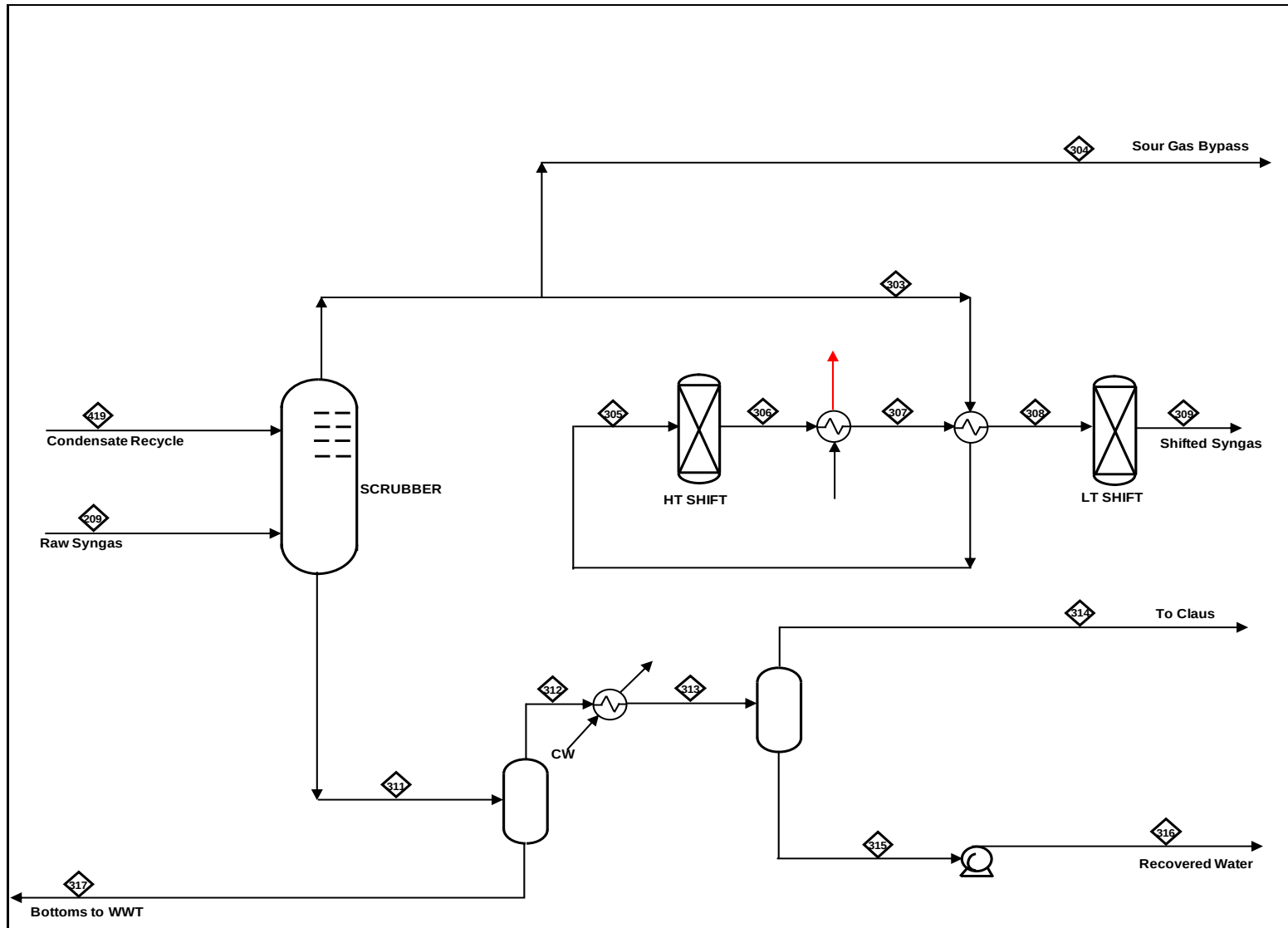


Table 5-4
Case 3c Plant – Wet Scrubber/Water Gas Shift Stream Table

STREAM	209	419	303	304	305	306	307	308	309	311	312	313	314	315	317
Description	Raw Syngas	Condensate Recycle	Syngas to Shift	Bypass Syngas	Hot Syngas	Hot 1st Shift Syngas	Warm 1st Shift Syngas	Cold 1st Shift Syngas	Hot 2nd Shift Syngas	Scrubber Bottoms	Scrubber Hot Vapor	Scrubber Cold Vapor	Scrubber Sour Gas	LP Recycle Water	Bottoms to WWT
Mole Flow (Vapor/Liquid)/lbmol/hr															
AR	795	0	524	270	524	524	524	524	524	1	1	1	1	0	0
CH4	353	0	233	120	233	233	233	233	233	0	0	0	0	0	0
CO	64212	2	42353	21841	42353	16096	16096	16096	8222	20	20	20	20	0	0
CO2	2002	27	1335	689	1335	27615	27615	27615	35490	5	5	5	5	0	0
COS	36	0	23	12	23	3	3	3	1	0	0	0	0	0	0
H2	25998	4	17149	8844	17149	43409	43409	43409	51282	9	9	9	9	0	0
H2O	66531	63740	51430	26522	51430	25148	25148	25148	17273	52318	9931	9931	2	9929	42387
H2S	311	4	205	106	205	225	225	225	227	4	4	4	3	1	0
N2	928	0	612	316	612	612	612	612	612	0	0	0	0	0	0
NH3	90	28	66	34	66	69	69	69	69	15	10	10	0	10	5
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	3	0	2	1	2	0	0	0	0	0	0	0	0	0	0
HCL	3	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	161262	63804	113933	58754	113933	113933	113933	113933	113933	52373	9981	9981	41	9940	42392
Total V/L Flow, lb/hr	3215573	1150135	2257561	1164189	2257561	2257577	2257577	2257577	2257581	943787.1	180085	180085	995	179090	763702
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	3215573	1150135	2257561	1164189	2257561	2257577	2257577	2257577	2257581	943787	180085	180085	995	179090	763702
Pressure, psia	899	915	899	899	894	887	878	869	861	894	51	46	46	46	51
Temperature, F	602	392	435	435	530	953	600	507	636	435	284	130	130	130	284

Figure 5-5
Case 3c Plant – Low Temperature Gas Cooling PFD

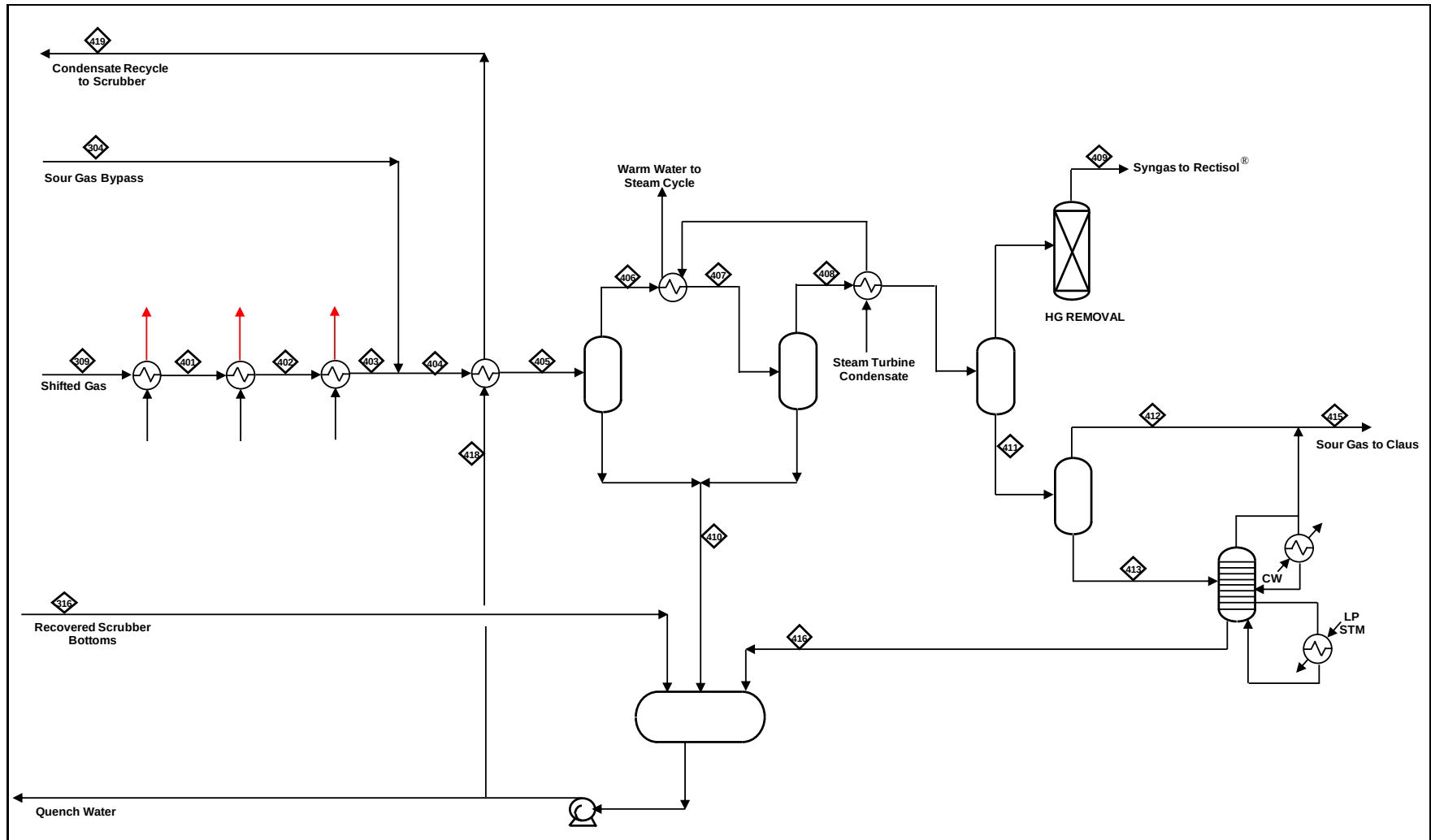


Table 5-5
Case 3c Plant – Low Temperature Gas Cooling Stream Table

STREAM	309	403	304	404	405	406	407	408	409	415	416	316	419	213
Description	Hot 2nd Shift Syngas	Cooled Shift Syngas	Bypass Syngas	Mixed Syngas	Mixed Syngas Shift	1st KO Drum Vapor	Cool 1st KO Drum Vapor	2nd KO KO Drum Vapor	Syngas to Rectisol®	Sour Gas	Sour Stripper Bottoms	HP Recycle Water	Condensate Recycle	Quench Water
Mole Flow (Vapor/Liquid)/lbmol/hr														
AR	524	524	270	794	794	794	794	794	794	0	0	0	0	0
CH4	233	233	120	353	353	353	353	353	353	0	0	0	0	0
CO	8222	8222	21841	30063	30063	30062	30062	30062	30062	0	0	0	2	0
CO2	35490	35490	689	36179	36179	36164	36164	36152	36149	3	0	0	27	0
COS	1	1	12	14	14	14	14	14	14	0	0	0	0	0
H2	51282	51282	8844	60126	60126	60123	60123	60122	60121	0	0	0	4	0
H2O	17273	17273	26522	43795	43795	28876	28876	14412	245	129	14038	9929	63740	64000
H2S	227	227	106	332	332	331	331	330	329	0	0	1	4	0
N2	612	612	316	927	927	927	927	927	927	0	0	0	0	0
NH3	69	69	34	103	103	95	95	85	0	85	0	10	28	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	1	1	1	1	1	1	1	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	113933	113933	58754	172687	172687	157740	157740	143251	128996	217	14039	9940	63804	64000
Total V/L Flow, lb/hr	2257581	2257581	1164189	3421771	3421771	3152137	3152137	2890814	2633987	3912.93	252914	179090	1150135	1152978
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	2257581	2257581	1164189	3421771	3421771	3152137	3152137	2890814	2633987	3913	252914	179090	1150135	1152978
Pressure, psia	861	846	899.0	846.0	843.0	837	834	834	815	61	74.16	46	915	980
Temperature, F	636	422	435	424	367	350	306	306	111	277	307	550	392	392

Figure 5-6
Case 3c Plant – Rectisol® AGR and CO₂ Compression PFD

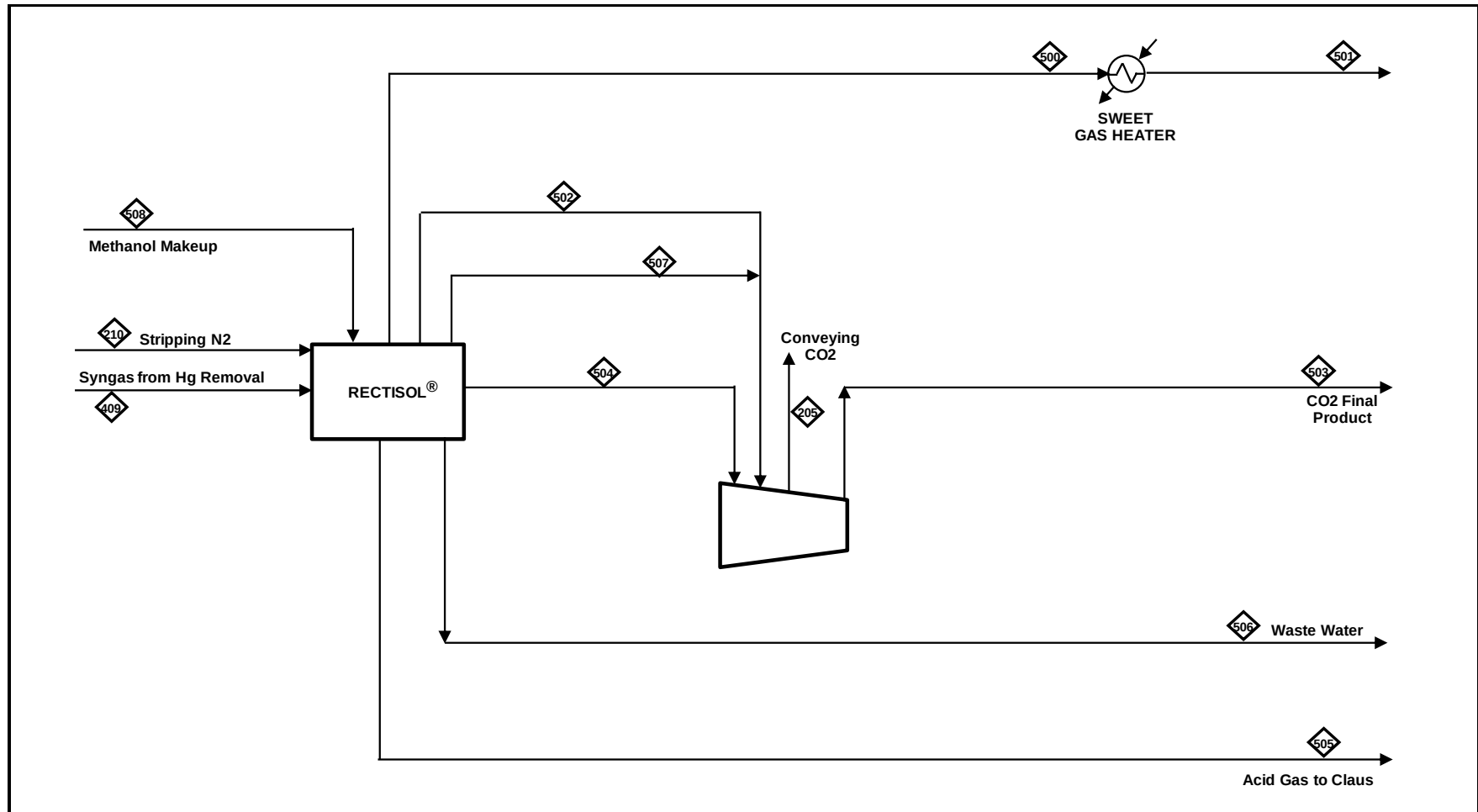


Table 5-6
Case 3c Plant – Rectisol® AGR and CO₂ Compression Stream Table

STREAM	409	210	500	504	502	503	205	505	506	507	508
Description	Syngas to Rectisol®	Stripping N ₂ to Rectisol®	Treated Gas to MeOH	LP CO ₂ Product Stream	MP CO ₂ Product Stream	CO ₂ Final Product	Convey CO ₂	Acid Gas to Claus	Water to Waste	Fuel Gas to CO ₂ Product	Makeup Rectisol®
Mole Flow (Vapor/Liquid)/lbmol/hr											
AR	794	1	790	0	2	4	0	0	0	3	0
CH ₄	353	0	344	0	5	8	1	0	0	4	0
CO	30062	0	29955	2	65	95	9	3	0	37	0
CO ₂	36149	0	3307	21921	9646	29215	2890	737	0	538	0
COS	14	0	0	0	0	0	0	14	0	0	0
H ₂	60121	0	60106	0	7	12	1	2	0	6	0
H ₂ O	245	0	0	0	0	0	0	0	245	0	0
H ₂ S	329	0	0	0	0	0	0	329	0	0	0
N ₂	927	586	892	618	1	563	56	3	0	1	0
NH ₃	0	0	0	0	0	0	0	0	0	0	0
O ₂	0	3	0	0	0	3	0	0	0	3	0
SO ₂	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0
HCN	1	0	0	0	0	0	0	1	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH ₃ OH	0	0	8	5	2	7	1	2	0	0	17
Ethanol -- C ₂ H ₅ OH	0	0	0	0	0	0	0	0	0	0	0
Propanol -- C ₃ H ₈ O	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C ₂ H ₆	0	0	0	0	0	0	0	0	0	0	0
Propane -- C ₃ H ₈	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C ₄ H ₁₀	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	128996	591	95402	22546	9729	29907	2959	1090	245	591	17
Total V/L Flow, lb/hr	2633987	16584	1168052	982280	426638	1304832	129095	44720	4423	25008	551
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	2633987	16584	1168052	982280	426638	1304832	129095	44720	4423	25008	551
Pressure, psia	815	75	785	19	42	2215	1100	28	36	75	40
Temperature, F	111	86	81	81	81	162	269	59	261	81	123

Figure 5-7
Case 3c Plant – Claus Plant PFD

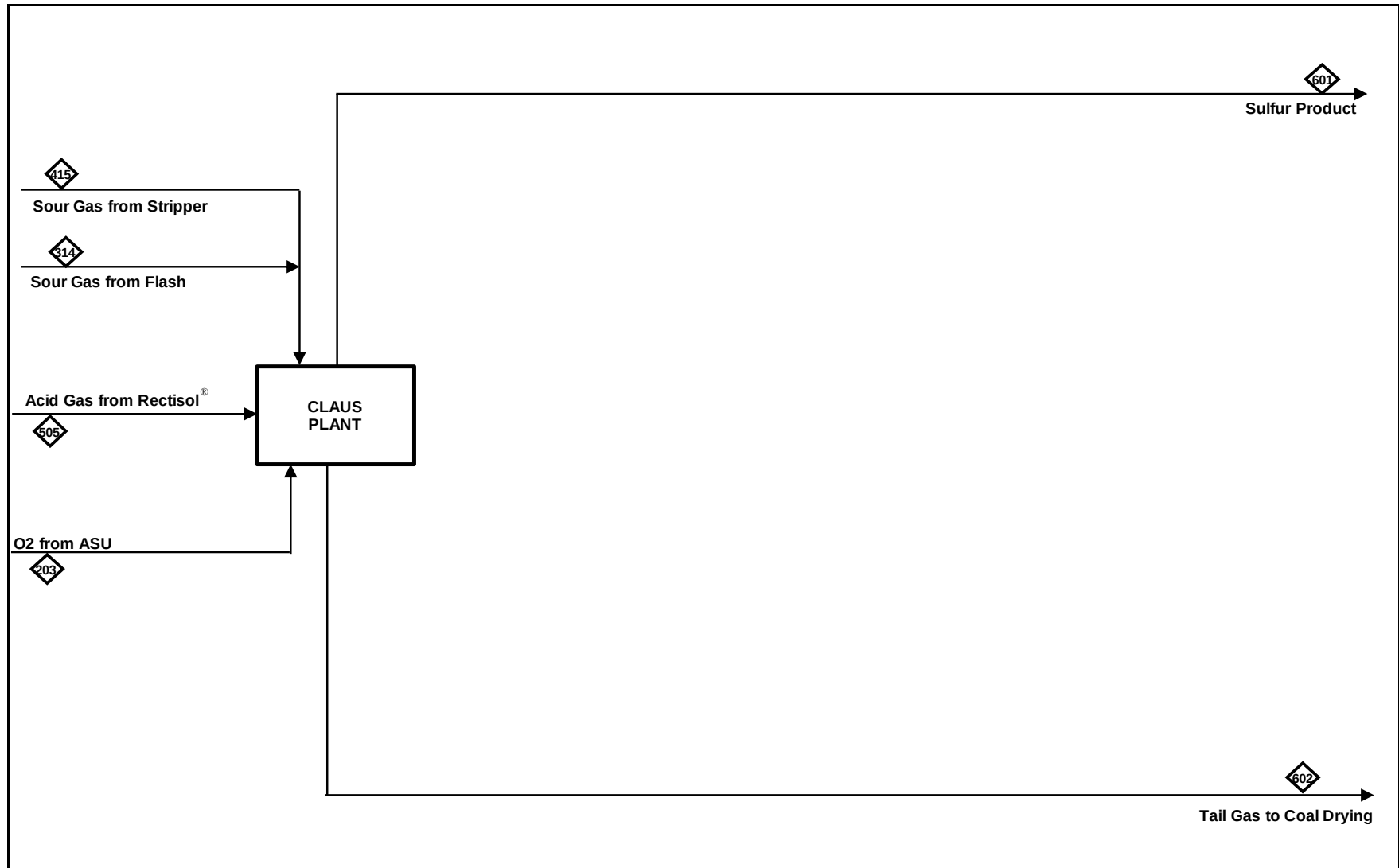


Table 5-7
Case 3c Plant – Claus Plant Stream Table

STREAM	505	415	314	203	601	602
Description	Acid Gas from Rectisol®	Sour Gas from Stripper	Sour Gas from Flash	Claus O2 from ASU	Sulfur Product	Tail Gas to Coal Drying
Mole Flow (Vapor/Liquid)/lbmol/hr						
AR	0	0	1	7	0	9
CH4	0	0	0	0	0	1
CO	3	0	20	0	0	23
CO2	737	3	5	0	0	757
COS	14	0	0	0	0	1
H2	2	0	9	0	0	10
H2O	0	129	2	0	0	588
H2S	329	0	3	0	0	4
N2	3	0	0	5	0	51
NH3	0	85	0	0	0	0
O2	0	0	0	237	0	0
SO2	0	0	0	0	0	2
S	0	0	0	0	340	0
HCN	1	0	0	0	0	1
HCL	0	0	0	0	0	0
Methanol -- CH3OH	2	0	0	0	0	2
Ethanol -- C2H5OH	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	1090	217	41	249	340	1448
Total V/L Flow, lb/hr	44720	3913	995	8014	10900	46743
Solids Mass Flow, lb/hr						
Coal	0	0	0	0	0	0
Ash	0	0	0	0	0	0
Carbon	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	44720	3913	995	8014	10900	46743
Pressure, psia	28	61	46	23	20	20
Temperature, F	59	277	130	72	320	320

Figure 5-8
Case 3c Plant – Methanol Synthesis Plant PFD

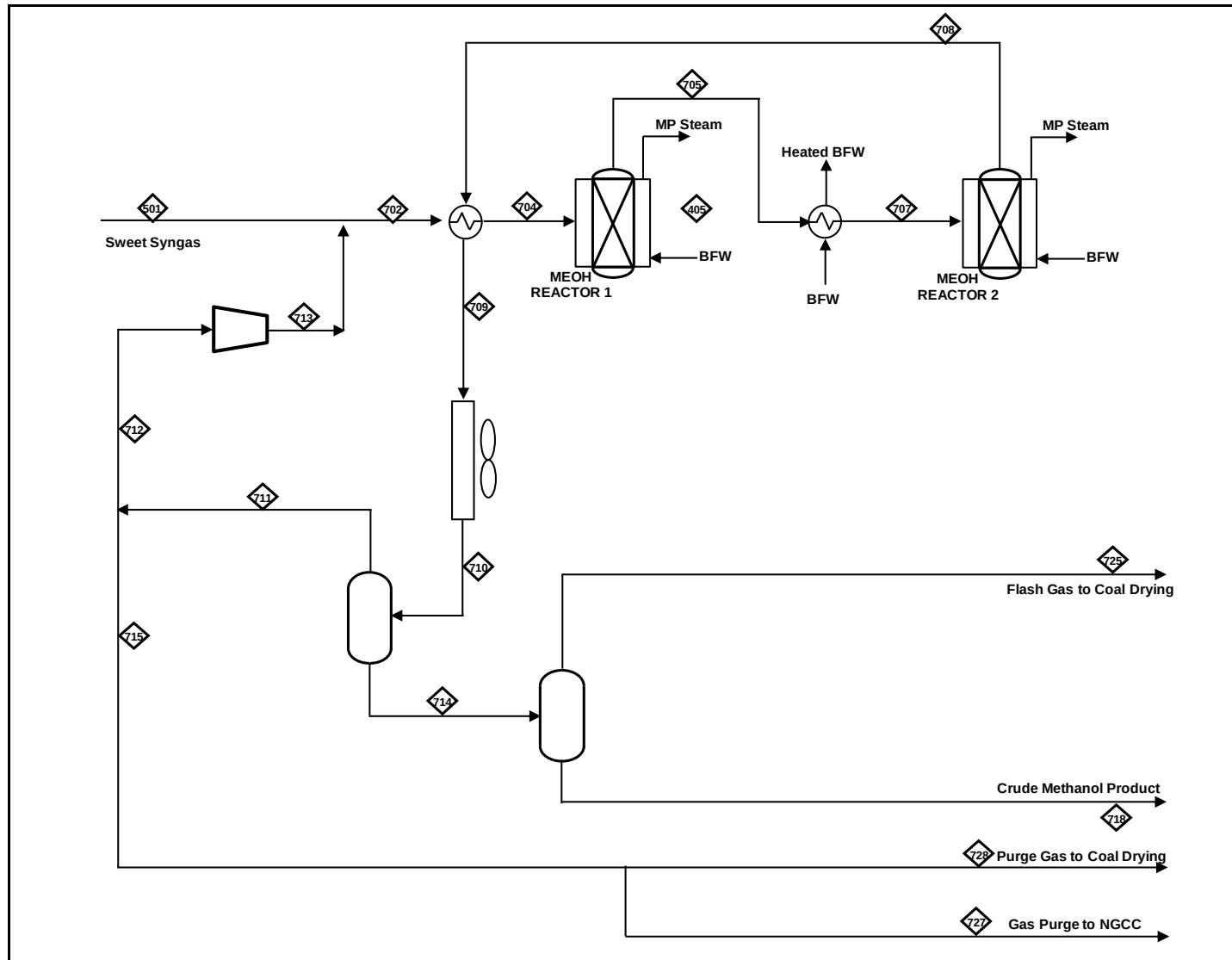


Table 5-8
Case 3c Plant – Methanol Synthesis Plant Stream Table

STREAM	501	704	705	707	708	709	710	711	712	713	714	718	725	727	728
Description	Reheated Syngas	MEOH Reactor 1 Feed	MEOH Reactor 1 Outlet	MEOH Reactor 2 Feed	MEOH Reactor 2 Outlet	Cooled Raw MeOH	AC Cool Raw MeOH	Flash Gas Ovhd	Recycle Syngas	HP Recycle Syngas	KO Drum Bottoms	Crude MeOH Product	MeOH Flash Gas	Fuel Gas to GT	MeOH Purge Gas
Mole Flow (Vapor/Liquid)/lbmol/hr															
AR	790	17190	17190	17190	17190	17190	17190	17083	16400	16400	107	2	105	120	564
CH4	344	7082	7082	7082	7082	7082	7082	7018	6738	6738	64	2	62	49	232
CO	29955	55062	35420	35420	26208	26208	26208	26152	25107	25107	56	0	55	183	863
CO2	3307	33403	32883	32883	33114	33114	33114	31351	30097	30097	1763	331	1432	219	1035
COS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	60106	96002	55159	55159	37425	37425	37425	37392	35896	35896	33	0	33	262	1234
H2O	0	11	577	577	348	348	348	11	11	11	336	335	2	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	892	20135	20135	20135	20135	20135	20135	20045	19243	19243	90	1	89	140	661
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	8	2753	22825	22825	31805	31805	31805	2859	2745	2745	28945	28542	404	20	94
Ethanol -- C2H5OH	0	3	45	45	46	46	46	3	3	3	43	42	0	0	0
Propanol -- C3H8O	0	0	2	2	2	2	2	0	0	0	2	2	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	95401.83	231641	191318	191318	173353	173353	173353	141915	136239	136239	31439	29258	2181	993	4683
Total V/L Flow, lb/hr	1168052	4658765	4658730	4658730	4658714	4658714	4658714	3636123	3490714	3490714	1022591	937317	85274	25453	119992
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	1168052	4658765	4658730	4658730	4658714	4658714	4658714	3636123	3490714	3490714	1022591	937317	85274	25453	119992
Pressure, psia	783	747	737	732.0	727	722	720	717	717	755	717	40	40	717	717
Temperature, F	125	400	475	400	430	229	130	130	130	141	130	118	118	130	130

Figure 5-9
Case 3c Plant – NGCC PFD

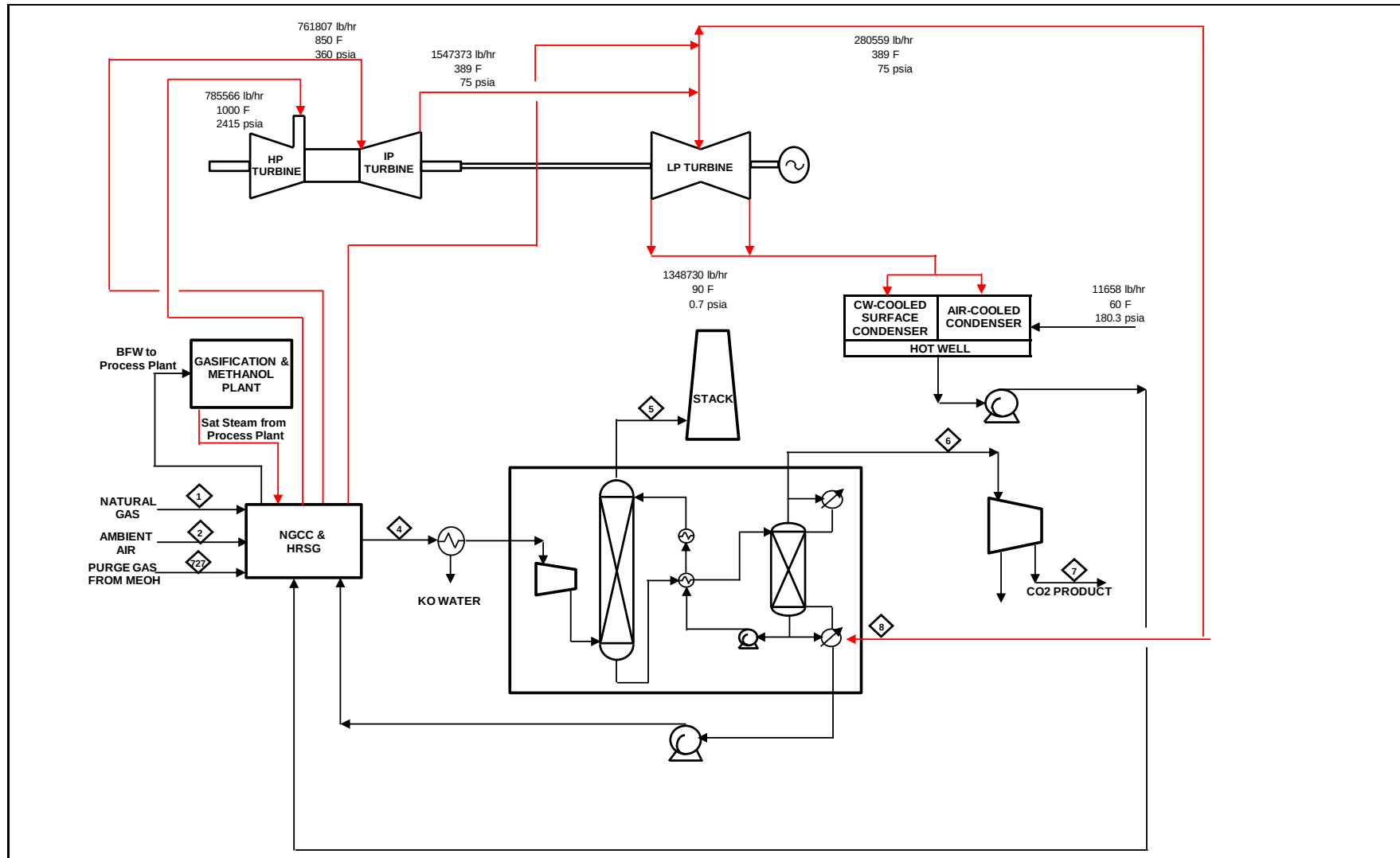


Table 5-9
Case 3c Plant – NGCC Stream Table

STREAM	1	2	4	5	6	7	8	727
Description	Ambient Air	Natural Gas	Flue Gas to MEA	Treated Gas	CO2 Out	CO2 Product	LP Steam	Fuel Gas to NG Boiler
Mole Flow (Vapor/Liquid)/lbmol/hr								
AR	1057	0	1177	1177	0	0	0	120
CH4	0	3473	0	0	0	0	0	69
CO	0	0	0	0	0	0	0	183
CO2	38	37	4395	440	3956	3956	0	219
COS	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	262
H2O	1128	0	9011	3594	45	0	15573	0
H2S	0	0	0	0	0	0	0	0
N2	88394	60	88594	88594	0	0	0	140
NH3	0	0	0	0	0	0	0	0
O2	23712	0	15761	15761	0	0	0	0
SO2	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	119	0	0	0	0	0	0
Propane -- C3H8	0	26	0	0	0	0	0	0
n-Butane -- C4H10	0	15	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	114330	3730	118938	109566	4000	3956	15573	993
Total V/L Flow, lb/hr	3299204	64622	3388952	3117277	174894	174088	280559	25115
Total Flow (Solids + V/L), lb/hr	3299204	64622	3388952	3117277	174894	174088	280559	25115
Pressure, psia	14.5	474.7	15	15	24	2215	260	176
Temperature, F	59	60	298	85	69	124	35	365

5.3 SPARING PHILOSOPHY

The sparing philosophy for the AR-provided equipment for the Case 3c plant is provided below. The design has:

- Five trains of gasification, including dry feed system, AR compact gasifier, cyclone and candle filter (5 x 25%)

With regards to the rest of the CTM plant's systems that warrant consideration in this study, there was no sparing information provided in the Crude Methanol Study.

5.4 PERFORMANCE RESULTS

The Nexant-modeled Case 3c plant consumes 18,339 tpd of PRB coal and 34.0 MMSCFD of natural gas at the Midwestern site and produces 936,766 lb/hr (10,198 mtpd) of crude methanol on a net basis. Overall performance for the Case 3c CTM plant is summarized in Table 5-10 which includes the plant's auxiliary power requirements. Loads that are unique to the AR gasification system are shown in bold and italicized.

Table 5-10
Case 3c Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case 3c
Gas Turbine Power	127,774
Steam Turbine Power	199,060
TOTAL POWER, kWe	326,834
CTM Plant Auxiliary Load Summary, kWe	
Coal Handling	8,585
Slag Handling	2,178
Incinerator Air Blower	4,477
Incinerator Recycle Blower	1,122
Air Separation Unit	152,266
Quench Water Pump	2,964
AR DSP	8,483
Scrubber Pumps	846
Flash Bottoms Pump	573
Rectisol® AGR Auxiliaries	31,510
Claus Plant Auxiliaries	235
CO ₂ Compressor Auxiliaries	52,408
Recycle Gas Compressor	3,244
Air Cooler Fans	1,769
Water Treatment	2,809
Miscellaneous BOP	5,000
Circulating Water Pumps	7,043
Ground Water Pumps	771
Cooling Tower Fans	4,592
SUBTOTAL CTM PLANT AUXILIARIES, kWe	290,876
NGCC Plant Auxiliary Load Summary, kWe	
Gas Turbine Auxiliaries	787
Steam Turbine Auxiliaries	100
Transformer Losses	1,793
Miscellaneous BOP	500
Air Cooled Condenser Fans	2,920
Condensate Pumps	563
Boiler Feed Water Pumps	10,984
Amine System Auxiliaries	4,141
NGCC CO ₂ Compression	6,411
SCR	16
SUBTOTAL POWER PLANT AUXILIARIES, kWe	28,214
TOTAL AUXILIARIES, kWe	319,041
NET PLANT POWER, kWe	7,744

Table 5-11 shows the carbon balance for the Case 3c plant. Carbon leaves the CTM process plant as unburned carbon in the slag, CO₂ in the dryer exhaust gas, ASU vent gas, crude methanol product, and the Rectisol® AGR CO₂ capture product. Additionally, carbon also leaves the power cycle as CO₂ in the stack and Econamine CO₂ capture product. Carbon in the crude methanol is considered to be product and not emissions. The carbon capture efficiency is defined as the amount of carbon in the product streams, which include the crude methanol and the dried and compressed CO₂ products from the Rectisol® AGR and Econamine CO₂ capture processes, relative to the amount of carbon in the coal and natural gas feedstock less carbon contained in the slag. For Case 3c, the carbon capture efficiency is 92.1% (same as for Case 3a).

Table 5-11
Case 3c Plant – Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	765,169	
ASU Air	511	
Coal Dryer Air	69	
Power Cycle Natural Gas	46,682	
Power Cycle Combustion Air	450	
ASU Vent		511
Coal Dryer Exhaust Gas		59,665
Carbon in Slag		0
Sulfur Product		0
Crude Methanol Product		347,703
Rectisol® AGR CO ₂ Product		352,217
Power Cycle CO ₂ Product		47,511
Power Cycle Exhaust Gas		5,279
Convergence Tolerance		-5
Total	812,881	812,881

Table 5-12 shows the sulfur balance for the Case 3c plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur emitted in the coal dryer exhaust gas, and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible. The net sulfur emissions include only the sulfur emitted in the coal dryer exhaust gas. Based on this, the net sulfur emissions for the plant are 0.040 lb of SO₂/MMBtu, which meet the sulfur emission limit in the BACT environmental design basis in Table 2-7. However, this sulfur emission exceeds the environmental target used in the companion IGCC studies performed on the advanced technologies. Actual permitting for the plant will also include the methanol process in addition to the gasification and power generation, which generally targets lower sulfur emissions. If the permit for this plant requires lower sulfur emissions, additional sulfur removal would be required for this plant.

Table 5-12
Case 3c Plant – Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	11,118	
Sulfur Product		10,901
Coal Dryer Exhaust Gas		216
CO ₂ Product		0
Convergence Tolerance		1
Total	11,118	11,118

Table 5-13 shows the overall water balance for the Case 3c plant. Raw water is obtained from groundwater (50%) and from municipal sources (50%). Some water is discharged by the processes as effluent suitable for internal consumption. This effluent is internally recycled and consumed by the slag handling process, as make up to the syngas scrubber, and as makeup to the cooling tower. Some water is discharged from the process to a permitted outfall. Waste water is discharged by the slag handling process, the Rectisol[®] unit, and as cooling tower blowdown.

Table 5-13
Case 3c Plant – Overall Water Balance

Water Use, gpm	Raw Water Withdrawal	Process Effluent Production for Internal Consumption	Internal Consumption	Process Water Discharge
Slag Handling	0	0	324	(324)
Quench Cooler	2,304	0	0	0
Syngas Scrubber Consumption	374	0	1,920	0
Syngas Scrubber Effluent	0	(1,883)	0	0
Syngas Cooling Knockout	0	(1,058)	0	0
Sour Water Stripper Effluent	0	(505)	0	0
Rectisol Unit	0	0	0	(9)
Steam Cycle Makeup	23	0	0	0
Steam Cycle Blowdown	0	(23)	0	0
Flue Gas Cooling Knockout	0	(193)	0	0
CO ₂ Compression Knockout	0	(2)	0	0
Cooling Tower Makeup	5,762	0	1,420	0
Cooling Tower Blowdown	0	0	0	(1,628)
Total	8,463	(3,664)	3,664	(1,961)

Positive values represent consumption while negative values represent production

5.5 EQUIPMENT LIST

The list in Table 5-14 below shows the Case 3b AR-related equipment that differs from the reference Case 3a reference plant.

Table 5-14
Case 3c Equipment List

ACCOUNT 2 COAL PREPARATION AND FEED
Subaccount 2.3 Dry Coal Injection System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	AR DSP Feed System	AR Proprietary dry solids pump with coal feed bins	1,000 tpd	15	0

ACCOUNT 4 GASIFIER, ASU AND ACCESSORIES
Subaccount 4.1 Gasifier & Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	Gasifier	Vertical, down-fired gasifier with multiple fuel injection ports and including water quench	3,000 tpd	5	0
4	Candle Filters	Ceramic	10,000 acfm	5	0
5	Syngas Cyclone	High efficiency	600,000 lb/hr	5	0

5.6 CAPITAL COST

Table 5-15 shows the cost breakdown of the Case 3c plant, consistent with the code of accounts format as expressed in the Crude Methanol Study. The accounts/subaccounts of interest for this study are:

- 2.3 Dry Coal Injection System,
- 4 Gasifier & Accessories,
- 5A Gas Cleanup & Piping and
- 5B CO₂ Removal and Compression

These are shown with more detail to include the various subaccounts and to provide more clarity as to wherein the major cost differences lie among the cases.

Table 5-16 shows the calculation and addition of owner's costs to determine the TOC, which is used to calculate the methanol RSP.

The estimated TOC of the Case 3c plant in 2011 dollars is \$4,857MM.

Table 5-15
Case 3c Total Plant Cost Summary

Case 3c: AR CTM Plant with Rectisol®-based AGR Low Rank Western Coal Baseline Study			Coal Feed, lb/hr Coal HHV, Btu/lb		1,528,260 8,564		Plant Size		10198 metric tons per day			
PRB												
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost \$	Eng'g CM H.O & Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect				Process	Project	\$	
	1 COAL & SORBENT HANDLING	\$35,251	\$6,321	\$27,325	\$0	\$0	\$68,897	\$6,253	\$0	\$15,030	\$90,180	
	2 COAL & SORBENT PREP & FEED											
	2.3 Dry Coal Injection System	\$134,811	\$0	\$88,517	\$0	\$0	\$223,328	\$19,235	\$44,666	\$57,446	\$344,674	
	2.x Other Coal & Sorbent Prep & Feed Systems	\$146,191	\$27,729	\$42,212	\$0	\$0	\$216,131	\$18,906	\$0	\$47,008	\$282,045	
	SUBTOTAL 2.	\$281,001	\$27,729	\$130,729	\$0	\$0	\$439,459	\$38,141	\$44,666	\$104,454	\$626,719	
	3 FEEDWATER & MISC BOP SYSTEMS	\$27,818	\$7,801	\$16,702	\$0	\$0	\$52,321	\$4,916	\$0	\$12,995	\$70,232	
	4 GASIFIER & ACCESSORIES											
	4.1 Gasifier, Syngas Cooler & Auxiliaries (AR)	\$88,269	\$44,134	\$133,274	\$0	\$0	\$265,677	\$23,821	\$66,419	\$53,387	\$409,304	
	4.3 ASU/Oxidant Compression	\$541,342	\$0	\$0	\$0	\$0	\$541,342	\$52,472	\$0	\$59,381	\$653,194	
	4.4 LT Heat Recovery & FG Saturation	\$76,905	\$0	\$34,490	\$0	\$0	\$111,395	\$10,872	\$0	\$24,453	\$146,720	
	4.6 Flare Stack System	\$0	\$1,421	\$647	\$0	\$0	\$2,068	\$198	\$0	\$453	\$2,719	
	4.9 Gasification Foundations	\$0	\$30,489	\$18,183	\$0	\$0	\$48,671	\$4,456	\$0	\$13,282	\$66,409	
	SUBTOTAL 4.	\$706,515	\$76,044	\$186,594	\$0	\$0	\$969,153	\$91,819	\$66,419	\$150,956	\$1,278,347	
	5A GAS CLEANUP & PIPING											
	5A.1 Rectisol® System	\$314,589	\$0	\$0	\$0	\$0	\$314,589	\$29,721	\$62,917	\$81,446	\$488,673	
	5A.2 Elemental Sulfur Plant	\$9,640	\$1,879	\$12,352	\$0	\$0	\$23,871	\$2,319	\$0	\$5,238	\$31,428	
	5A.3 Mercury Removal	\$7,303	\$0	\$5,520	\$0	\$0	\$12,822	\$1,238	\$641	\$2,940	\$17,642	
	5A.4 Shift Reactors	\$11,008	\$0	\$4,400	\$0	\$0	\$15,408	\$1,477	\$0	\$3,377	\$20,263	
	5A.5 Syngas Compressor	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
	5A.6 Blowback Gas Systems	\$6,453	\$1,086	\$612	\$0	\$0	\$8,150	\$773	\$0	\$1,785	\$10,708	
	5A.7 Fuel Gas Piping	\$0	\$2,904	\$1,900	\$0	\$0	\$4,804	\$445	\$0	\$1,050	\$6,299	
	5A.9 HGCU Foundations	\$0	\$2,543	\$1,713	\$0	\$0	\$4,256	\$391	\$0	\$1,394	\$6,041	
	SUBTOTAL 5.	\$348,993	\$8,411	\$26,497	\$0	\$0	\$383,901	\$36,364	\$63,558	\$97,229	\$581,053	
	5B CO2 REMOVAL & COMPRESSION											
	5B.1 NGCC CO2 Removal System	\$23,800	\$0	\$7,172	\$0	\$0	\$30,972	\$2,584	\$6,194	\$7,951	\$47,701	
	5B.2 CO2 Compression & Drying (AACRP & NGCC)	\$61,147	\$0	\$23,051	\$0	\$0	\$84,198	\$8,067	\$0	\$18,453	\$110,718	
	5B.3 CO2 Compression & Drying (NGCC)	\$10,509	\$0	\$4,023	\$0	\$0	\$14,532	\$1,215	\$0	\$3,149	\$18,896	
	SUBTOTAL 5B.	\$95,456	\$0	\$34,246	\$0	\$0	\$129,702	\$11,866	\$6,194	\$29,553	\$177,315	
	5C METHANOL PRODUCTION	\$132,749	\$55,850	\$111,700	\$0	\$0	\$300,299	\$30,030	\$0	\$66,066	\$396,395	
	6 COMBUSTION TURBINE/ACCESSORIES	\$47,397	\$394	\$3,814	\$0	\$0	\$51,605	\$11,171	\$11,909	\$14,852	\$89,537	
	7 HRSG, DUCTING & STACK	\$25,164	\$703	\$5,977	\$0	\$0	\$31,844	\$2,644	\$0	\$3,468	\$37,956	
	8 STEAM TURBINE GENERATOR	\$81,770	\$895	\$20,740	\$0	\$0	\$103,405	\$9,290	\$0	\$21,949	\$134,643	
	9 COOLING WATER SYSTEM	\$17,205	\$18,871	\$14,046	\$0	\$0	\$50,122	\$4,637	\$0	\$11,142	\$65,901	
	10 ASH/SPENT SORBENT HANDLING SYS	\$77,771	\$3,478	\$73,220	\$0	\$0	\$154,469	\$14,945	\$0	\$17,687	\$187,101	
	11 ACCESSORY ELECTRIC PLANT	\$35,447	\$21,187	\$37,318	\$0	\$0	\$93,952	\$8,178	\$0	\$20,358	\$122,488	
	12 INSTRUMENTATION & CONTROL	\$12,231	\$3,697	\$10,258	\$0	\$0	\$26,187	\$2,351	\$1,309	\$5,348	\$35,195	
	13 IMPROVEMENTS TO SITE	\$5,786	\$2,843	\$15,142	\$0	\$0	\$23,770	\$2,347	\$0	\$7,835	\$33,952	
	14 BUILDINGS & STRUCTURES	\$0	\$7,656	\$10,622	\$0	\$0	\$18,278	\$1,652	\$0	\$3,235	\$23,165	
CALCULATED TOTAL COST		\$1,930,554	\$241,879	\$724,930	\$0	\$0	\$2,897,363	\$276,605	\$194,055	\$582,157	\$3,950,180	

Table 5-16
Case 3c Total Plant Cost Summary

Owner's Costs	\$ x \$1,000
Preproduction Costs	
6 months All Labor	\$28,311
1 Month Maintenance Materials	\$5,287
1 Month Non-Fuel Consumables	\$884
1 Month Waste Disposal	\$1,154
25% of 1 Months Fuel Cost at 100% CF	\$6,734
2% of TPC	\$79,004
Total	\$121,373
Inventory Capital	
60 day supply of fuel at 100% CF	\$53,131
60 day supply of non-fuel consumables at 100% CF	\$1,245
0.5% of TPC (spare parts)	\$19,751
Total	\$74,127
Initial Cost for Catalyst and Chemicals	\$10,944
Land	\$900
Other Owner's Cost	\$592,527
Financing Costs	\$106,655
Total Owner's Costs	\$906,527
Total Overnight Costs (TOC)	\$4,856,707

5.7 OPERATING COSTS

Table 5-17 shows the OPEX breakdown for the Case 3c plant.

Table 5-17
Case 3c Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Case 3c - AR Gasifier with Rectisol® AGR CTM Plant				
Plant Size (mtpd Methanol)	10,198				
Primary/Secondary Fuel:	PRB				
Design/Construction	5 years		Book Life (yrs):	20	
TPC (Plant Cost) Year	June 2011		TPI Year:	2016	
Capacity Factor (%)	90		CO2 Captured (TPD)	15461	
OPERATING & MAINTENANCE LABOR					
Operating Labor					
Operating Labor Rate (base):	\$39.70		\$/hr		
Operating Labor Burden:	30.00		% of base		
Labor Overhead Charge	25.00		% of labor		
Operating Labor Requirements per Shift	units/mod		Total Plant		
Skilled Operator	2.0		2.0		
Operator	10.0		10.0		
Foreman	1.0		1.0		
Lab Tech's etc	3.0		3.0		
TOTAL Operating Jobs	16.0		16.0		
					<u>Annual Cost</u>
					\$
Annual Operating Labor Cost	\$7,233,658				
Maintenance Labor Cost	\$38,063,936				
Administration & Support Labor	\$11,324,398				
Property Taxes and Insurance	\$79,003,603				
TOTAL FIXED OPERATING COSTS					\$135,625,596
VARIABLE OPERATING COSTS					
Maintenance Material Cost					\$57,095,904
<u>Consumables</u>	<u>Consumption</u>	<u>Unit</u>	<u>Initial Fill</u>		
	<u>Initial</u>	<u>/Day</u>	<u>Cost</u>	<u>Cost</u>	
Water(/1000 gallons)	0	6,093	1.67	\$0	\$3,350,470
Chemicals					
MU & WT Chem (lb)	0	36305	0.27	\$0	\$3,194,345
Carbon (Hg Removal) (lb)	249237	380	1.63	\$406,256	\$203,290
MEA Solvent (ton)	161	0.22	3751.70	\$605,845	\$272,164
Water Gas Shift Catalyst (ft3)	10061	6.89	771.99	\$7,766,827	\$1,748,165
Methanol Synthesis Catalyst	4050	3.69	534.68	\$2,165,562	\$648,948
Claus Catalyst (ft3)	w/equip	1.94	203.15	\$0	\$129,440
Subtotal Chemicals				\$10,944,490	\$6,196,353
Other					
Supplemental Electricity (MWh consumed)	0	0	62.33	\$0	\$0
Gases, N2 etc./100scf)	0	0	0.00	\$0	\$0
LP Steam (/1000 lbs)	0	0	0.00	\$0	\$0
Subtotal Other				\$0	\$0
Waste Disposal:					
Spent Mercury Catalyst (lb)	0	380	0.65	\$0	\$81,067
Flyash (ton)	0	0	0.00	\$0	\$0
Slag (ton)	0	1502	25.11	\$0	\$12,385,523
Subtotal Waste Disposal				\$0	\$12,466,590
By-products & Emissions					
Sulfur (tons)	0	130.8	0.00	\$0	\$0
Supplemental Electricity (MWh generated)	0	186	-59.59	\$0	-\$3,638,065
Subtotal By-Products				\$0	-\$3,638,065
TOTAL VARIABLE OPERATING COSTS				\$10,944,490	\$75,471,252
Coal (tons)	0	18,339	36.57	\$0	\$220,312,342
Natural Gas (MMBtu)	0	35,051	6.13	\$0	\$70,582,488

5.8 METHANOL PRODUCT REQUIRED SELLING PRICE

Table 5-18 shows a summary of the power output, CAPEX, OPEX, and methanol product RSP for the Case 3c plant. The Case 3c plant methanol RSP is estimated to be \$1.24/gal under the loan guarantee finance structure and \$1.45/gal under the commercial fuels finance structure.

Table 5-18
Plant Performance and Economic Summary

Case	Case 3c	
CAPEX, \$MM		
Total Installed Cost (TIC)	\$2,897	
Total Plant Cost (TPC)	\$3,950	
Total Overnight Cost (TOC)	\$4,857	
OPEX, \$MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC _{fix})	\$135.6	
Variable Operating Cost, less Fuel (OC _{var})	\$87.9	
Coal Feedstock (OC _{coal})	\$244.8	
Natural Gas Feedstock (OC _{NG})	\$78.4	
Import/(Export) Power (OC _{Power})	(\$4.0)	
Total OPEX	\$542.7	
Plant Output		
Crude Methanol Product, tons per year	4,103,035	
Net Power Output, MWe	7.74	
Required Selling Price^A		
Excluding CO₂ TS&M^{BE}, \$/ton	359.5	422.7
Including CO₂ TS&M^{BE}, \$/ton	375.1	438.3
RSP Component Details (\$/gal)		
Capital ^B	0.74	0.95
Fixed O&M	0.12	
Variable O&M	0.07	
Coal	0.20	
Natural Gas	0.06	
Power	(0.00)	
CO ₂ TS&M	0.05	
RSP^B Total (\$/gal)	1.24	1.45
Costs of CO₂ Captured^{B,C} (\$/tonne)	18.7	21.0
Costs of CO₂ Avoided^{B,D} (\$/tonne)	29.7	32.0

^A Capacity factor assumed to be 90 percent

^B Values shown are for two financial structures

The first (lower value) is based on the loan guarantee finance structure

The second (higher value) is based on the commercial fuels finance structure

^C Excludes CO₂ TS&M

^D Includes CO₂ TS&M

^E Based on 301.73 gallons/ton

6.1 PROCESS OVERVIEW

The final case to be evaluated, Case 3d Shell Gasifier with RTI WDP CTM plant, has the same configuration as Case 3b, with the exception that the AR compact gasifier is replaced by the original Shell gasifier from the Case 3a reference plant. The Case 3d plant is equipped with an NGCC to generate enough power to meet the plant's auxiliary demands such that no power needs to be imported, and any excess power is exported for extra revenue. The Case 3d plant is designed to capture CO₂ with a carbon capture efficiency of more than 90% of the carbon in the coal and natural gas. The Case 3d plant operates with an annual on-stream CF of 90%.

Case 3d, the standalone RTI WDP/AFWGS/AACRP case, is established so that it can be compared with Case 3b to identify any synergistic advantages in integrating both the AR and RTI systems in a CTM plant.

The Case 3d plant has the following characteristics that differentiate it from the Case 3a reference plant:

- Syngas leaving the gasifier that has been quenched and filtered enters RTI's WDP for sulfur removal at about 755°F instead of going through a low temperature scrubber as in Case 3a.
- After sulfur removal in the WDP, a portion of the treated syngas in Case 3d then enters the WGS process. Instead of the sour shift WGS process used in Case 3a, the Case 3d AFWGS process consists of fixed-bed reactors (using commercial high-temperature sweet WGS catalyst) combined in such a manner as to significantly reduce the overall steam consumption and reactor capital cost while still meeting the manufacturer's steam to CO recommendations. These reactors are operated at standard temperatures for commercial high-temperature sweet WGS processes, but at a higher inlet temperature than commercial sour WGS process.
- After the syngas is shifted to a 2:1 H₂/CO ratio and cooled, it enters the AACRP for CO₂ capture. Unlike Rectisol[®], which is a physical solvent, activated methyldiethanolamine is a chemical solvent. As ~99.9% of the sulfur compounds have been removed by the WDP process upstream already, the amine process only has to remove CO₂ and is less complicated than the two-stage Rectisol[®] setup. However, the AACRP unit does also capture ~99% of any residual sulfur left in the syngas following WDP along with the CO₂, resulting in an overall system total sulfur reduction of >99.99% (sub-ppmv total sulfur in the final cleaned syngas).
- The DSRP replaces the Claus process in Case 3a. In the DSRP, sulfur leaving the WDP process in the form of SO₂ is reduced by a slipstream of shifted syngas, forming elemental sulfur, H₂S and COS. The elemental sulfur is condensed while the remaining H₂S and COS are re-oxidized in the presence of air to SO₂. The SO₂ is then removed in a

lime scrubber, forming gypsum ($\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$). This combined approach for sulfur capture results in very low net SO_2 emissions.

The coal feed rate is expected to be the same as Case 3a since Case 3d also utilizes the Shell gasifier operating under the same conditions. It therefore has the same cold gas efficiency and generates the same amount of raw gasifier syngas that is eventually converted to methanol product.

6.2 CASE 3d PROCESS DESCRIPTION

The overall BFD for the Case 3d plant is shown in Figure 6-1, with the accompanying stream flows shown in Table 6-1. Additional descriptions of the CTM plant's various processes are provided below. To better visualize the different unit operations in the Case 3d plant, simplified PFDs of the various plant processes are depicted in Figures 6-2 through 6-8. Tables 6-2 through 6-8 provide the model-generated process data for the numbered streams referenced in the PFDs.

Coal Preparation and Drying

Same as Case 3a. Figure 6-2 depicts the coal drying process while Table 6-2 presents the mass balances.

Coal Feed System

Same as Case 3a

Air Separation Unit

Same as Case 3a except the required oxygen is reduced as DSRP does not require oxygen like the Claus process in Case 3a.

Shell Gasifier

Same as Case 3a. The coal feed system, ASU, Shell gasification and quench processes are shown in Figure 6-3 and their mass balances are presented in Table 6-3.

Dry Solids Removal

Same description as Case 3a. The solids removal process is shown in Figure 6-3.

Unlike in Case 3a, the syngas scrubber is eliminated in Case 3d as the syngas leaving the raw syngas coolers at around 755°F enters the RTI WDP directly without further quench cooling.

Warm Syngas Cleanup Process

The Case 3d warm syngas cleanup process consists of five major system components: RTI WDP, RTI DSRP, AFWGS, LTGC, and the AACRP unit. These are described in greater detail below.

Warm Gas Desulfurization Process

Same as Case 3b except at lower pressure

Direct Sulfur Reduction Process

Same as Case 3b except at lower pressure

Advanced Fixed-Bed Water Gas Shift

Same as Case 3b except at lower pressure. The RTI WDP, DSRP and AFWGS processes are shown in Figure 6-4 and the mass balances are presented in Table 6-4.

Low Temperature Gas Cooling and Mercury Removal

Same as Case 3b except at lower pressure. The LTGC and mercury removal processes are shown in Figure 6-5 and the mass balances presented in Table 6-5.

Activated Amine CO₂ Removal Process

Same as Case 3b except the AACRP absorber operates at lower pressure.

CO₂ Compression and Dehydration

Same as Case 3b. The AACRP and CO₂ compression process are shown in Figure 6-6, while the material balance is shown in Table 6-6. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table.

Methanol Reactor and Synthesis Loop

Same as Case 3a. The methanol synthesis process is shown in Figure 6-7, while the material balance is shown in Table 6-7. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table.

NGCC

Same as Case 3a. The PFD for the NGCC section is shown in Figure 6-8, while the material balance is shown in Table 6-8. The stream numbers labeled on the PFD correspond to the stream numbers in the material balance table.

Power Cycle Flue Gas Post Combustion CO₂ Capture and Compression

Same as Case 3b

Figure 6-1
Case 3d Plant – Overall BFD

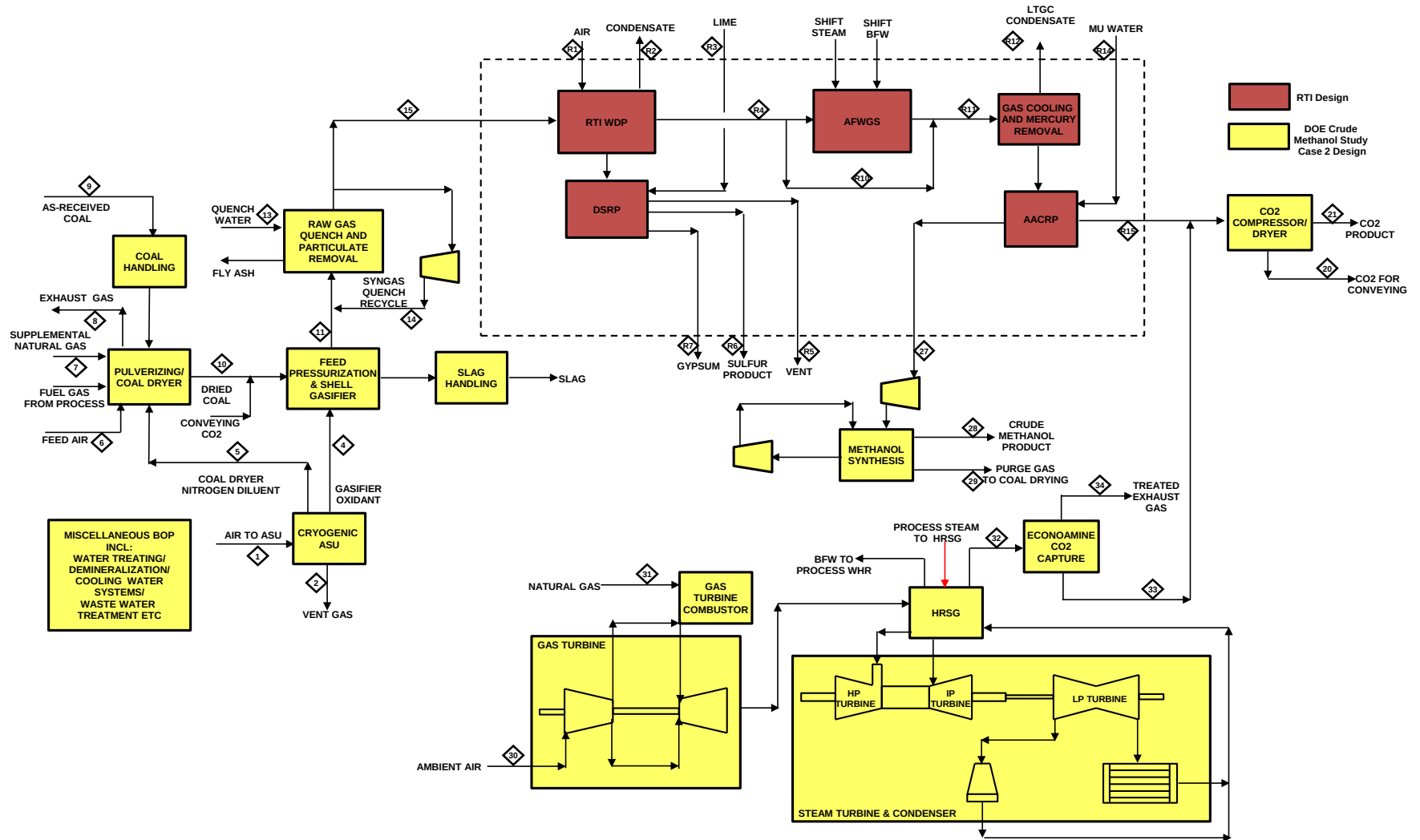


Table 6-1
Case 3d Plant – Overall Stream Table

Description	Air to ASU	ASU Vent Gas	Gasifier Oxidant	Nitrogen Diluent to Dryer	Air Feed to Coal Dryer	Supplemental NG Feed	Coal Dryer Exhaust	As-Received Coal	Dried Coal	Raw Syngas	Slag	Quench Water	Cool Syngas Recycle	Syngas to RTI WDP
Stream No.	1	2	4	5	6	7	8	9	10	11	12	13	14	15
V-L Mole Fraction														
Ar	0.0092	0.0234	0.0300	0.0023	0.0092	0	0.0087	0	0	0.0088	0	0	0.0054	0.0054
CH4	0	0	0	0	0	0.9310	0	0	0	0.0000	0	0	0.0000	0.0000
CO	0	0	0	0	0	0	0	0	0	0.6201	0	0	0.3786	0.3786
CO2	0.0003	0.0056	0	0	0.0003	0	0	0	0	0.0648	0	0	0.0396	0.0396
COS	0	0	0	0	0	0	0	0	0	0.0005	0	0	0.0003	0.0003
H2	0	0	0	0	0	0	0	0	0	0.2304	0	0	0.1407	0.1407
H2O	0.0099	0.1653	0	0	0	0	0.1446	0	0	0.0625	0	1.0000	0.4276	0.4276
H2S	0	0	0	0	0	0	0	0	0	0.0030	0	0	0.0018	0.0018
N2	0.7732	0.6762	0.0200	0.9921	0.7732	0.0160	0.8082	0	0	0.0097	0	0	0.0060	0.0060
NH3	0	0	0	0	0	0	0	0	0	0.0000	0	0	0	0
O2	0.2074	0.1295	0.9500	0.0054	0.2074	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0.0000	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0.0320	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0.0070	0	0	0	0	0	0	0	0
n-Butane	0	0	0	0	0	0.0040	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0	0	1.0000	0	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	151743	8924	31278	111542	18337	130	156104	0	0	106198	0	67741	135448	173939
V-L Flowrate, lb/hr	4378850	243127	1005820	3129914	529151	2252	4237613	0	0	2397107	0	1220373	2816973	3617480
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	1618190	1277850	0	136541	0	0	0
Temperature, F	59	64	292	70	59	80	176	59	176	2649		400	475	765
Pressure, psia	14.7	16.4	711.0	18.0	14.7	20.0	14.3	14.7	14.3	650.0	650.0	685.0	650.0	605.3

Table 6-1 (cont'd)
Case 3d Plant – Overall Stream Table

Description	Air to RTI WDP	Comp Air KO Water	Lime	Sweet Syngas	DSRP Vent Gas	Sulfur Product	Gypsum	Bypass Syngas	Syngas to Cooling	Cooled Syngas KO Water	Syngas to aMDEA Unit	Makeup Wash Water	aMDEA CO2 Out
Stream No.	R1	R2	R3	R4	R5	R6	R7	R10	R11	R12	R13	R14	R15
V-L Mole Fraction													
Ar	0.0092	0	0	0.0054	0.0134	0	0	0.0054	0.0044	0	0.0070	0	0
CH4	0	0	0	0.0000	0	0	0	0.0000	0.0000	0	0.0000	0	0
CO	0	0	0	0.3760	0	0	0	0.3760	0.1396	0	0.2213	0	0.0006
CO2	0.0003	0	0	0.0432	0.0668	0	0.0081	0.0432	0.1991	0.0001	0.3155	0	0.9099
COS	0	0	0	0.0000	0	0	0	0.0000	0.0000	0	0	0	0
H2	0	0	0	0.1441	0	0	0	0.1441	0.2805	0	0.4447	0	0.0016
H2O	0.0099	1.0000	0.9737	0.4235	0.0002	0	0.9678	0.4235	0.3701	0.9998	0.0016	1.0000	0.0880
H2S	0	0	0	0.0000	0	0	0	0.0000	0.0000	0	0	0	0
N2	0.7732	0	0	0.0077	0.9192	0	0.0003	0.0077	0.0062	0	0.0099	0	0
NH3	0	0	0	0	0	0	0	0	0	0.0001	0	0	0
O2	0.2074	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	1.0000	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0.0263	0	0	0	0.0010	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0.0228	0	0	0	0	0	0
Methanol	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0
Propanol	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane	0	0	0	0	0	0	0	0	0	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.000	1.000	1.000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	4875	32	9076	176380	3779	138	10043	62897	218324	80594	135617	4208	43795
V-L Flowrate, lb/hr	140677	570	176897	3666696	110498	4413	218914	1307544	4422350	1452222	2924566	75804	1823984
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature, F	59	80	75	685	103	299	173	685	844	95	95	122	122
Pressure, psia	14.7	13.0	13.0	594.0	470.0	14.7	412.0	594.0	569.0	549.0	534.0	531.0	20.7

Table 6-1 (cont'd)
Case 3d Plant – Overall Stream Table

Description	Convey'g CO2	CO2 Product	Treated Syngas to MeOH	Crude Methanol Product	MeOH Synthesis Purge Gas	Air to Gas Turbine	Natural Gas Feed	NGCC Flue Gas Exhaust	CO2 to Comp- ression	Treated NGCC Exhaust
Stream No.	20	21	27	28	29	30	31	32	33	34
V-L Mole Fraction										
Ar	0	0	0.0098	0.0001	0.1262	0.0092	0	0.0089	0	0.0097
CH4	0	0	0.0001	0	0.0006	0	0.9310	0	0	0
CO	0.0006	0.0006	0.3123	0	0.1284	0	0	0	0	0
CO2	0.9976	0.9978	0.0306	0.0111	0.2079	0.0003	0.0100	0.0349	0.9880	0.0038
COS	0	0	0	0	0	0	0	0	0	0
H2	0.0017	0.0015	0.6273	0	0.3313	0	0	0	0	0
H2O	0	0	0.0060	0.0181	0.0001	0.0099	0	0.0761	0.0120	0.0332
H2S	0	0	0	0	0	0	0	0	0	0
N2	0	0	0.0140	0.0001	0.1860	0.7732	0.0160	0.7473	0	0.8096
NH3	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0.2074	0	0.1328	0	0.1438
SO2	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0
HCl	0	0	0	0	0	0	0	0	0	0
Ca(OH)2	0	0	0	0	0	0	0	0	0	0
CaSO.2H2O	0	0	0	0	0	0	0	0	0	0
Methanol	0	0	0	0.9698	0.0195	0	0	0	0	0
Ethanol	0	0	0	0.0009	0	0	0	0	0	0
Propanol	0	0	0	0.0000	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0.032	0	0	0
Propane	0	0	0	0	0	0	0.007	0	0	0
n-Butane	0	0	0	0	0	0	0.004	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate, lbmol/hr	5691	37988	95002	29055	6525	114330	3930	118374	3764	109265
V-L Flowrate, lb/hr	250001	1669025	1163802	927811	158575	3299204	68087	3367302	164499	3106521
Solids Flow Rate, lb/hr	0	0	0	0	0	0	0	0	0	0
Temperature, F	237	162	139	119	130	59	60	277	69	85
Pressure, psia	770.0	2215.0	521.0	40	717	14.5	474.7	14.9	24	15.0

Figure 6-2
Case 3d Plant – Coal Drying BFD

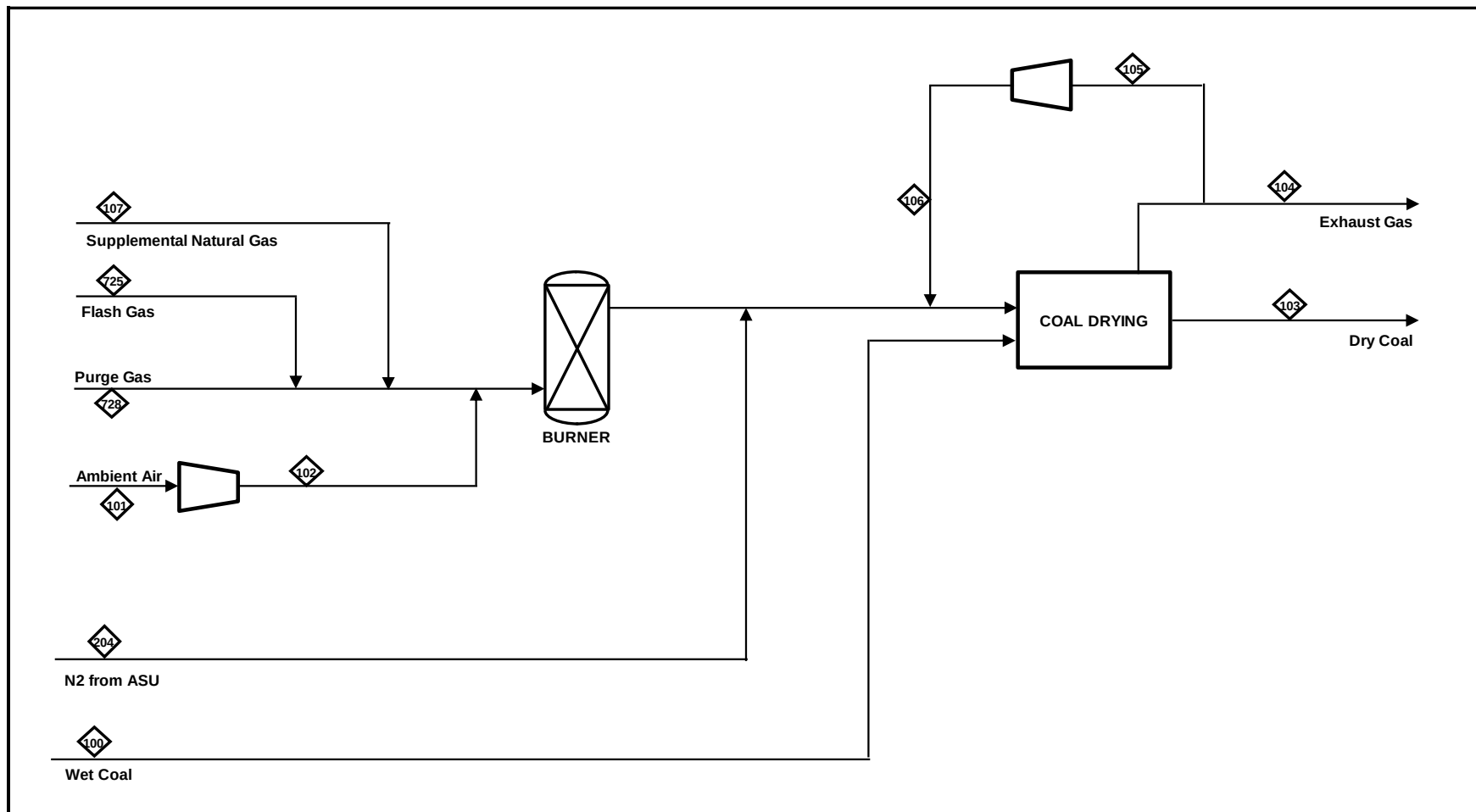


Table 6-2
Case 3d Plant – Coal Drying Stream Table

STREAM	100	101	102	103	104	105	106	107	204	725	728
Description	Wet Coal	Ambient Air	Feed Air	Dry Coal	Exhaust Gas	LP Recycle Gas	HP Recycle Gas	Supplemental Nat Gas	Dilution Nitrogen	MeOH Flash Gas	MeOH Purge Gas
Mole Flow (Vapor/Liquid)/lbmol/hr											
AR	0	170	170	0	1355	730	730	0	257	106	823
CH4	0	0	0	0	0	0	0	121	0	1	4
CO	0	0	0	0	0	0	0	0	0	37	838
CO2	0	6	6	0	4171	2246	2246	1	0	1294	1357
COS	0	0	0	0	0	0	0	0	0	0	0.00472
H2	0	0	0	0	0	0	0	0	0	39	2161
H2O	0	181	181	0	22568	12152	12152	0	22	2	1
H2S	0	0	0	0	0	0	0	0	0	0	0.00129
N2	0	14177	14177	0	126165.6	67935	67935	2	110661	112	1214
NH3	0	0	0	0	0	0	0	0	0	0	0
O2	0	3803	3803	0	1844	993	993	0	602	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0	0	371	127
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0.2	0.1
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	4	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	1	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	1	0	0	0
Total V/L Flow, lbmol/hr	0	18337	18337	0	156104	84056	84056	130	111542	1963	6525
Total V/L Flow, lb/hr	0	529151	529151	0	4237613	2281793	2281793	2252	3129914	77390	158575
Solids Mass Flow, lb/hr	1618190	0	0	1277850	0	0	0	0	0	0	0
Coal	1618190	0	0	1277850	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	1618190	529151	529151	1277850	4237613	2281793	2281793	2252	3129914	77390	158575
Pressure, psia	14.7	14.7	17	14.3	14.3	14.3	16	20	18	40	717
Temperature, F		59	90		176	176	203	80	70	119	130

Figure 6-3
Case 3d Plant – ASU/Coal Gasification PFD

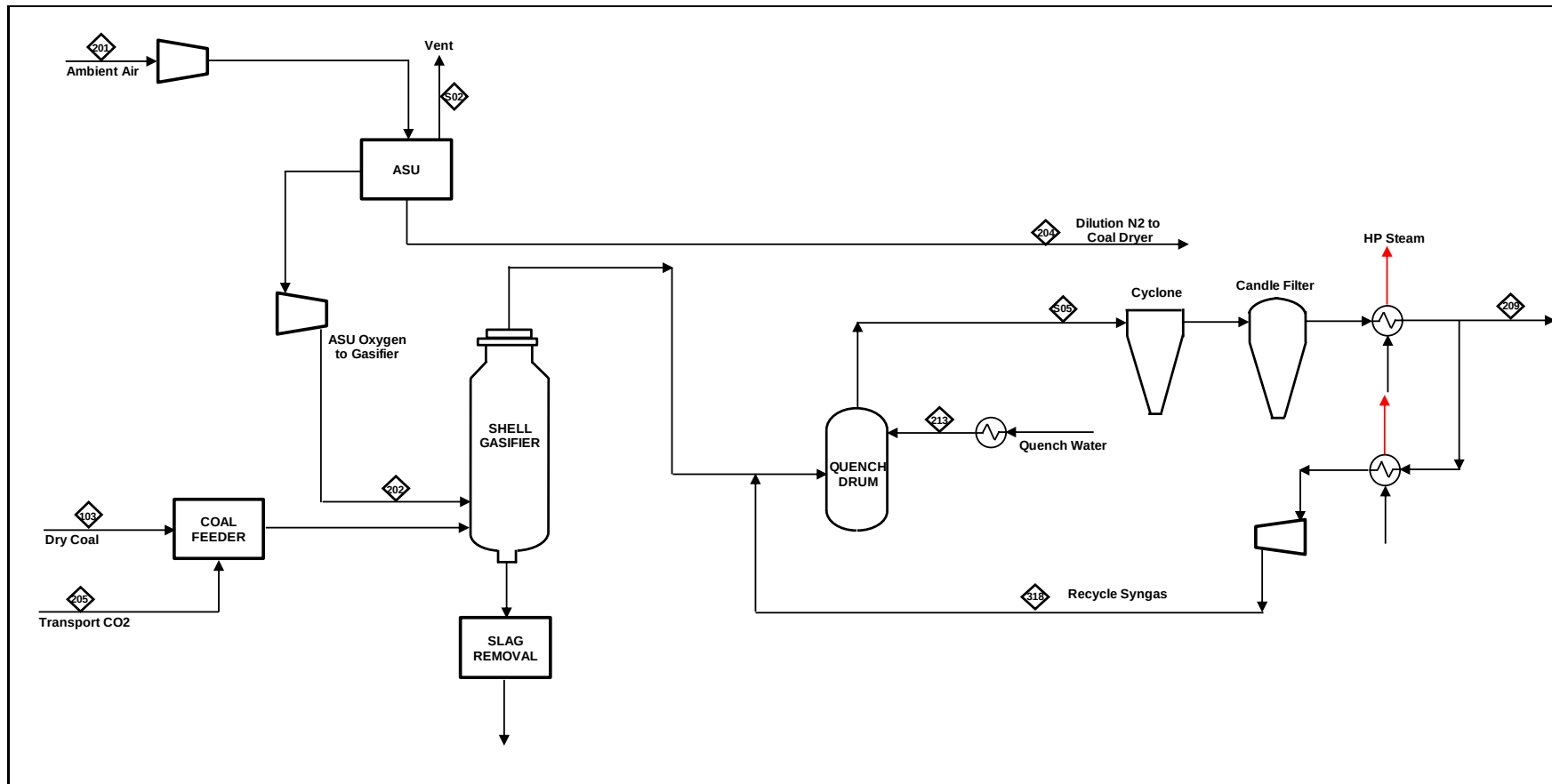


Table 6-3
Case 3d Plant – ASU/Coal Gasification Stream Table

STREAM	201	202	203	204	205	103	318	213	214	209
Description	Ambient Air	Oxygen to Gasifier	Oxygen to Claus	N2 to Coal Dryer	Conveying CO2	Dry Coal	Recycle Syngas	Quench Water	Slag Out	Raw Cooled Syngas
Mole Flow (Vapor/Liquid)/lbmol/hr										
AR	1403	938	0	257	0	0	731	0	0	938
CH4	0	0	0	0	0	0	4	0	0	5
CO	0	0	0	0	4	0	51283	0	0	65856
CO2	50	0	0	0	5678	0	5362	0	0	6885
COS	0	0	0	0	0	0	40	0	0	51
H2	0	0	0	0	10	0	19054	0	0	24468
H2O	1498	0	0	22	0	0	57916	67741	0	74375
H2S	0	0	0	0	0	0	246	0	0	316
N2	117321	626	0	110661	0	0	806	0	0	1035
NH3	0	0	0	0	0	0	3.584597	0	0	5
O2	31472	29714	0	602	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	1
HCL	0	0	0	0	0	0	3	0	0	3
Methanol -- CH3OH	0	0	0	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	151743	31278	0	111542	5691	0	135448	67741	0	173939
Total V/L Flow, lb/hr	4378850	1005820	0	3129914	250001	0	2816973	1220373	0	3617480
Solids Mass Flow, lb/hr	0	0	0	0	0	1277850	0	0	136541	0
Coal	0	0	0	0	0	1277850	0	0	0	0
Ash	0	0	0	0	0	0	0	0	132490	0
Carbon	0	0	0	0	0	0	0	0	4051	0
Total Flow (Solids + V/L), lb/hr	4378850	1005820	0	3129914	250001	1277850	2816973	1220373	136541	3617480
Pressure, psia	14.7	711	23.2	18.0	770.0	14.3	650	685	650	605.3
Temperature, F	59	292	72	70	237		475	400		765

Figure 6-4
Case 3d Plant – RTI WDP/DSRP/AFWGS PFD

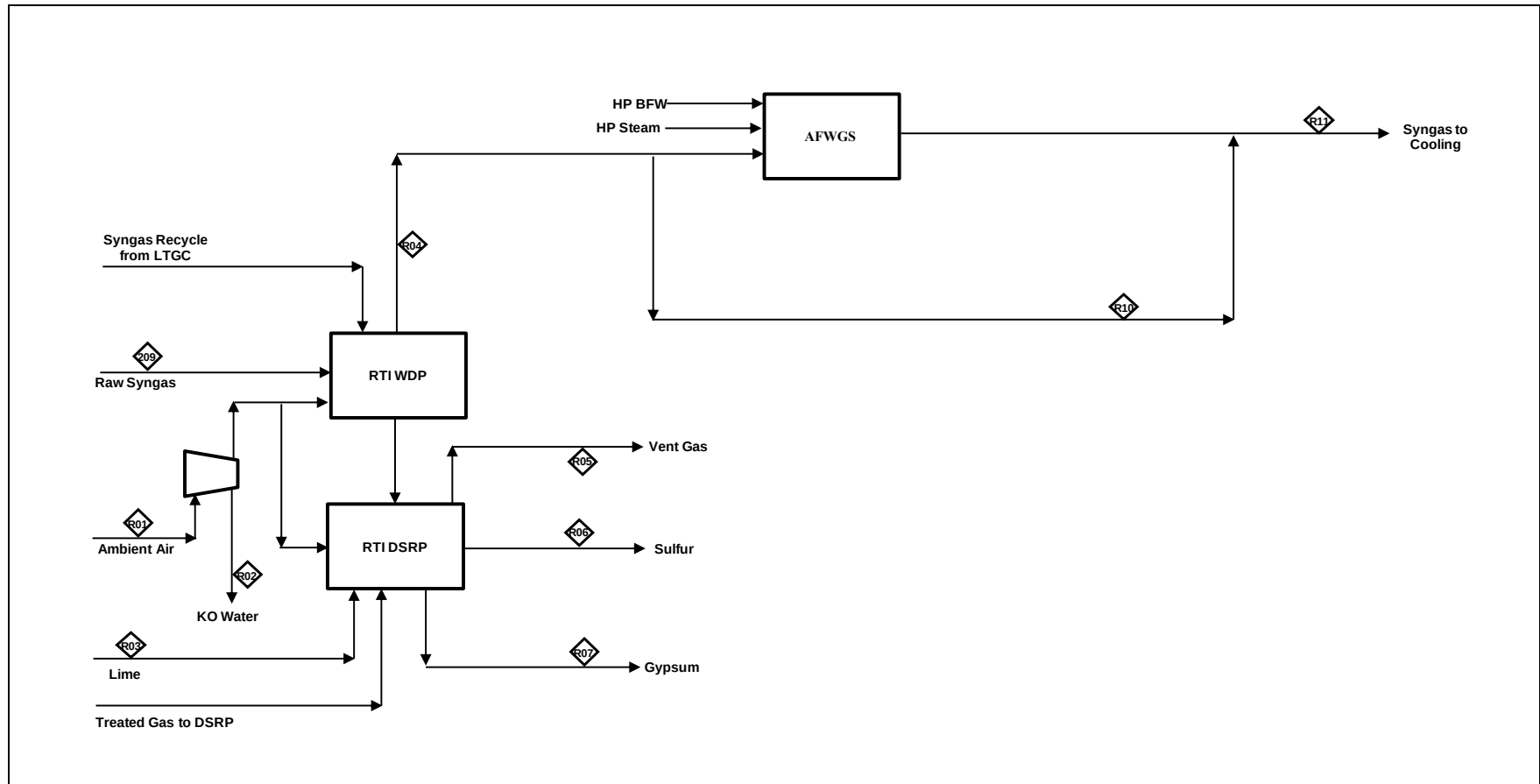


Table 6-4
Case 3b Plant – RTI WDP/DSRP/AFWGS Stream Table

STREAM	209	R01	R02	R03	R04	R05	R06	R07	R10	R11
Description	Raw Syngas	Ambient Air	Comp KO Water	Lime	Treated Gas to Shift	DSRP Vent Gas	Sulfur Product	Gypsum Product	Bypass Syngas	Syngas to LT Cooling
Mole Flow (Vapor/Liquid)/lbmol/hr										
AR	938	45	0	0	957	50	0	0	341	957
CH4	5	0	0	0	5	0	0	0	2	5
CO	65856	0	0	0	66324	0	0	0	23651	30486
CO2	6885	2	0	0	7623	253	0	82	2718	43461
COS	51	0	0	0	0	0	0	0	0	0
H2	24468	0	0	0	25408	0	0	0	9060	61246
H2O	74375	48	32	8837	74693	1	0	9720	26636	80799
H2S	316	0	0	0	1	0	0	0	0	1
N2	1035	3769	0	0	1363	3473	0	3	486	1363
NH3	5	0	0	0	5	0	0	0	2	5
O2	0	1011	0	0	0	1	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	138	0	0	0
HCN	1	0	0	0	0	0	0	0	0	0
HCL	3	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0
Lime	0	0	0	239	0	0	0	10	0	0
Gypsum	0	0	0	0	0	0	0	229	0	0
Total V/L Flow, lbmol/hr	173939	4875	32	9076	176380	3779	138	10043	62897	218324
Total V/L Flow, lb/hr	3617480	140677.5	569.5217	176896.8	3666696	110498	4412.519	218913.8	1307544	4422350
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	3617480	140677	570	176897	3666696	110498	4413	218914	1307544	4422350
Pressure, psia	605	14.7	13.0	13.0	594.0	470	14.7	412	594	569
Temperature, F	765	59	80	75	685	103	299	173	685	844

Figure 6-5
Case 3d Plant – LTGC PFD

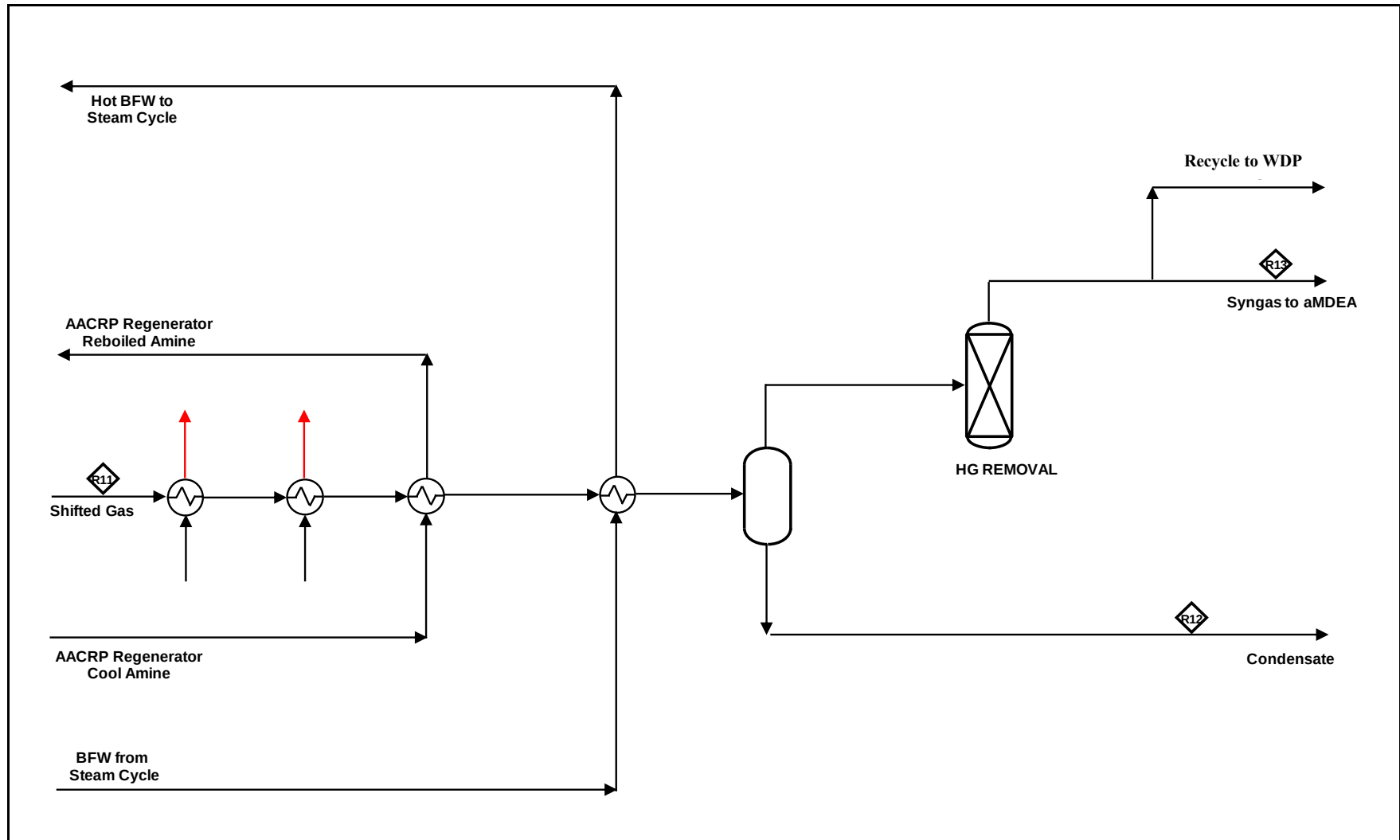


Table 6-5
Case 3d Plant – LTGC Stream Table

STREAM	R11	R12	R13
Description	Shifted Syngas	KO Water	Cooled Syngas
Mole Flow (Vapor/Liquid)/lbmol/hr			
AR	957	0	943
CH4	5	0	5
CO	30486	0	30018
CO2	43461	12	42783
COS	0	0	0
H2	61246	0	60306
H2O	80799	80577	219
H2S	1	0	1
N2	1363	0	1342
NH3	5	5	0
O2	0	0	0
SO2	0	0	0
S	0	0	0
HCN	0	0	0
HCL	0	0	0
Methanol -- CH3OH	0	0	0
Ethanol -- C2H5OH	0	0	0
Propanol -- C3H8O	0	0	0
Ethane -- C2H6	0	0	0
Propane -- C3H8	0	0	0
n-Butane -- C4H10	0	0	0
Total V/L Flow, lbmol/hr	218324	80594	135617
Total V/L Flow, lb/hr	4422350	1452222	2924566
Solids Mass Flow, lb/hr	0	0	0
Coal	0	0	0
Ash	0	0	0
Carbon	0	0	0
Total Flow (Solids + V/L), lb/hr	4422350	1452222	2924566
Pressure, psia	569	549	534
Temperature, F	844	95	95

Figure 6-6
Case 3d Plant – AACRP and CO₂ Compression PFD

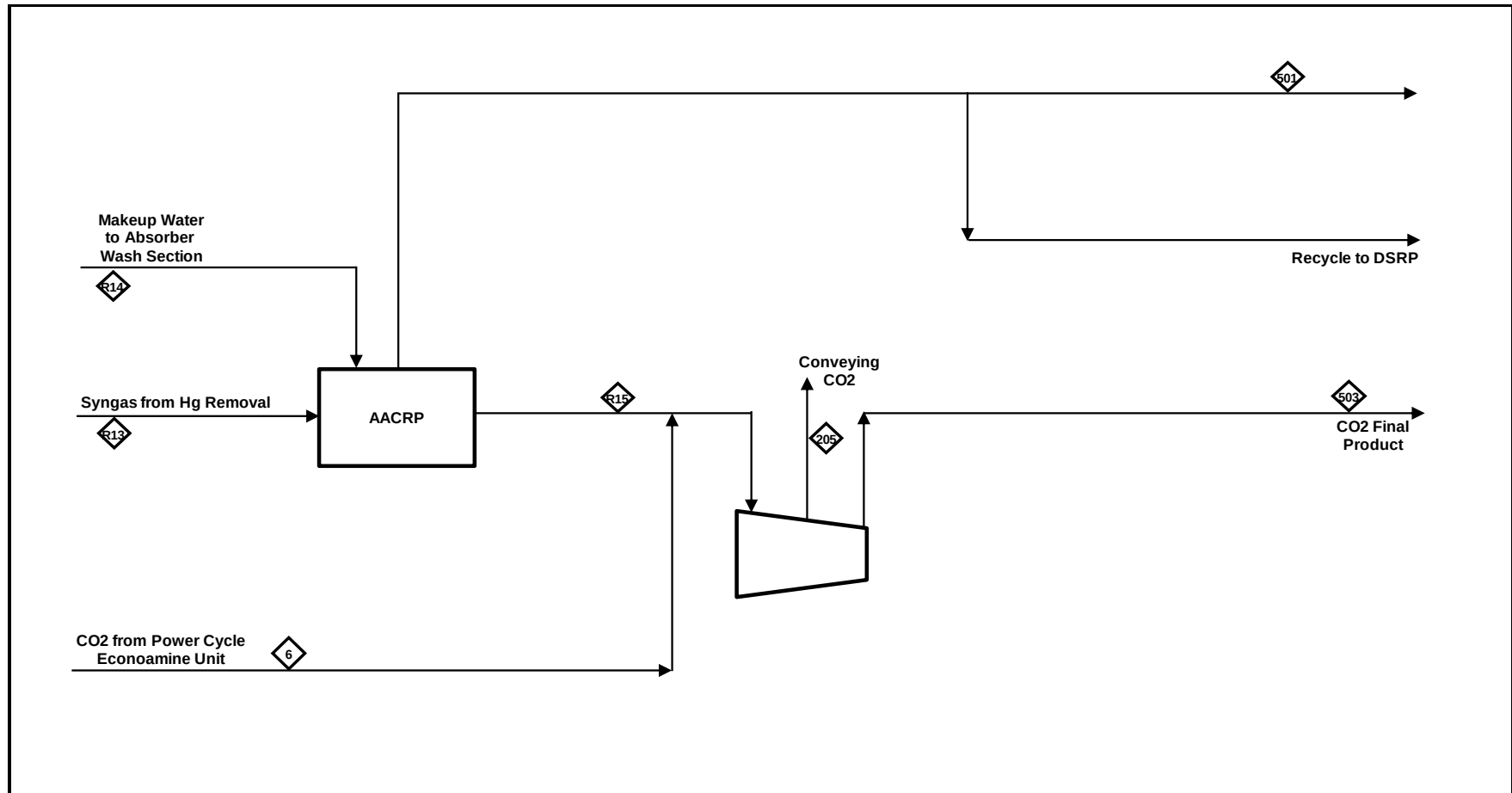


Table 6-6
Case 3d Plant – AACRP and CO₂ Compression Stream Table

STREAM	R13	R14	R15	501	6	205	503
Description	Syngas to AACRP	Makeup Wash Water	AACRP CO2 Product	Syngas to MeOH Synthesis	CO2 from MEA Unit	Convey CO2	CO2 Final Product
Mole Flow (Vapor/Liquid)/lbmol/hr							
AR	957	0	1	931	0	0	1
CH4	5	0	0	5	0	0	0
CO	30486	0	25	29672	0	4	21
CO2	43450	0	39847	2905	3736	5678	37906
COS	0	0	0	0	0	0	0
H2	61246	0	68	59594	0	10	59
H2O	222	4208	3852	568	45	0	0
H2S	1	0	1	0	0	0	1
N2	1363	0	1	1327	0	0	1
NH3	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0
Ethane -- C2H6	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	137729	4208	43795	95002	3782	5691	37988
Total V/L Flow, lb/hr	2970128	75804	1823984	1163802	165256	250001	1669025
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	2970128	75804	1823984	1163802	165256	250001	1669025
Pressure, psia	534	531	20.7	521	24	770	2215
Temperature, F	95	122	122	139	69	237	162

Figure 6-7
Case 3d Plant – Methanol Synthesis Plant PFD

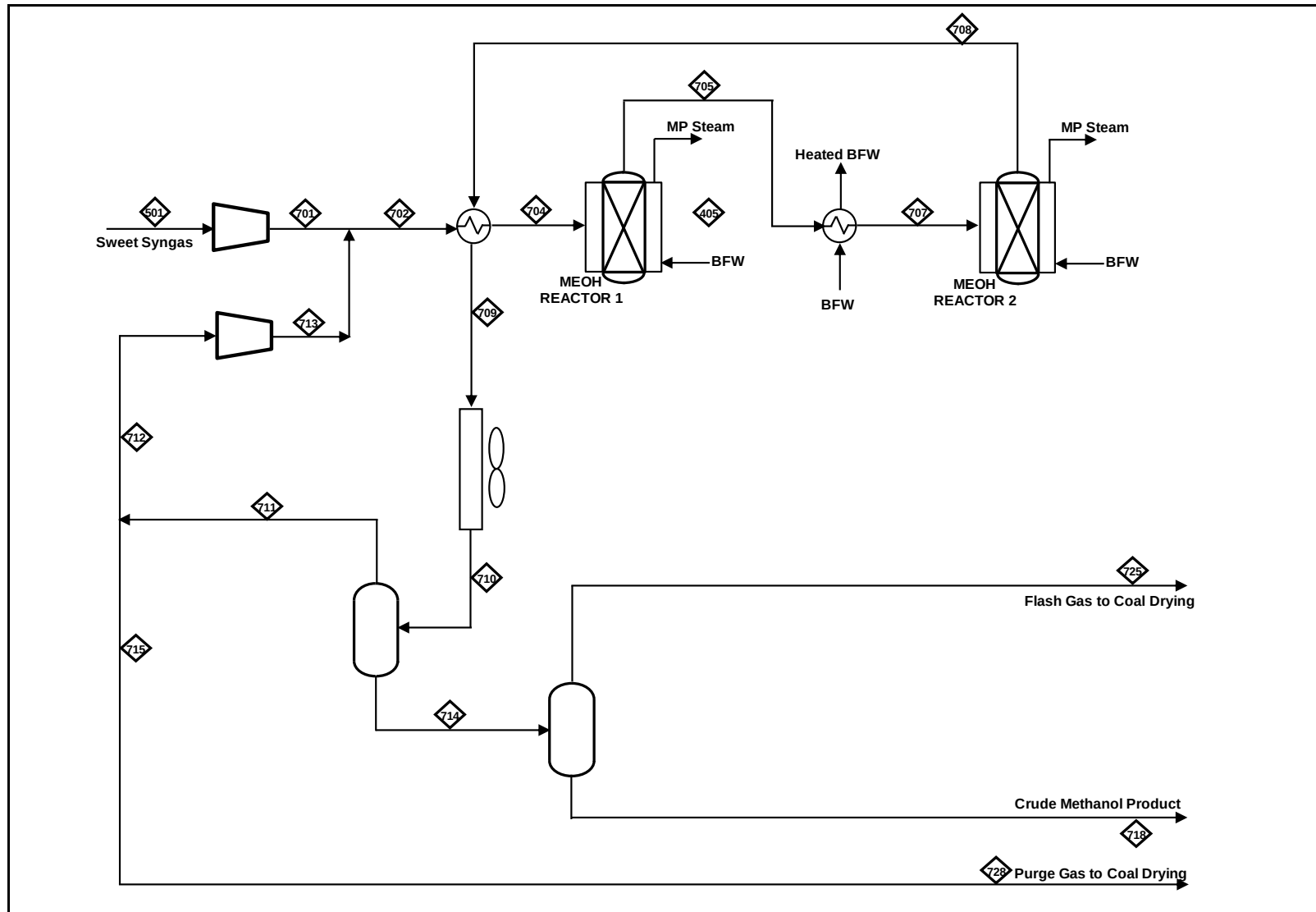


Table 6-7
Case 3d Plant – Methanol Synthesis Plant Stream Table

STREAM	501	704	705	707	708	709	710	711	712	713	714	718	725	728
Description	Reheated Syngas	MEOH Reactor 1 Feed	MEOH Reactor 1 Outlet	MEOH Reactor 2 Feed	MEOH Reactor 2 Outlet	Cooled Raw MeOH	AC Cool Raw MeOH	Flash Gas Ovhd	Recycle Syngas	HP Recycle Syngas	KO Drum Bottoms	Crude MeOH Product	MeOH Flash Gas	MeOH Purge Gas
Mole Flow (Vapor/Liquid)/lbmol/hr														
AR	931	16572	16572	16572	16572	16572	16572	16464	15641	15641	108	2	106	823
CH4	5	85	85	85	85	85	85	84	80	80	1	0	1	4
CO	29672	45589	25631	25631	16790	16790	16790	16753	15916	15916	37	0	37	838
CO2	2905	28679	28541	28541	28747	28747	28747	27130	25774	25774	1617	323	1294	1357
COS	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	59594	100660	60333	60333	43269	43269	43269	43229	41067	41067	40	0	39	2161
H2O	568	583	754	754	543	543	543	16	15	15	527	525	2	1
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	1327	24392	24392	24392	24392	24392	24392	24279	23065	23065	113	2	112	1214
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Methanol -- CH3OH	0	2417	22446	22446	31092	31092	31092	2544	2417	2417	28548	28176	371	127
Ethanol -- C2H5OH	0	1	33	33	28	28	28	2	1	1	26	26	0	0
Propanol -- C3H8O	0	0	1	1	1	1	1	0	0	0	1	1	0	0
Ethane -- C2H6	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane -- C3H8	0	0	0	0	0	0	0	0	0	0	0	0	0	0
n-Butane -- C4H10	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total V/L Flow, lbmol/hr	95002.48	218979	178789	178789	161520	161520	161520	130502	123977	123977	31017	29055	1963	6525
Total V/L Flow, lb/hr	1163802	4176747	4176712	4176712	4176697	4176697	4176697	3171496	3012945	3012945	1005201	927811	77389.9	158575
Solids Mass Flow, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Flow (Solids + V/L), lb/hr	1163802	4176747	4176712	4176712	4176697	4176697	4176697	3171496	3012945	3012945	1005201	927811	77390	158575
Pressure, psia	521	747	737	732.0	727	722	720	717	717	755	717	40	40	717
Temperature, F	138.501	400	475	400	430	244	130	130	130	141	130	119	119	130

Figure 6-8
Case 3d Plant -- NGCC PFD

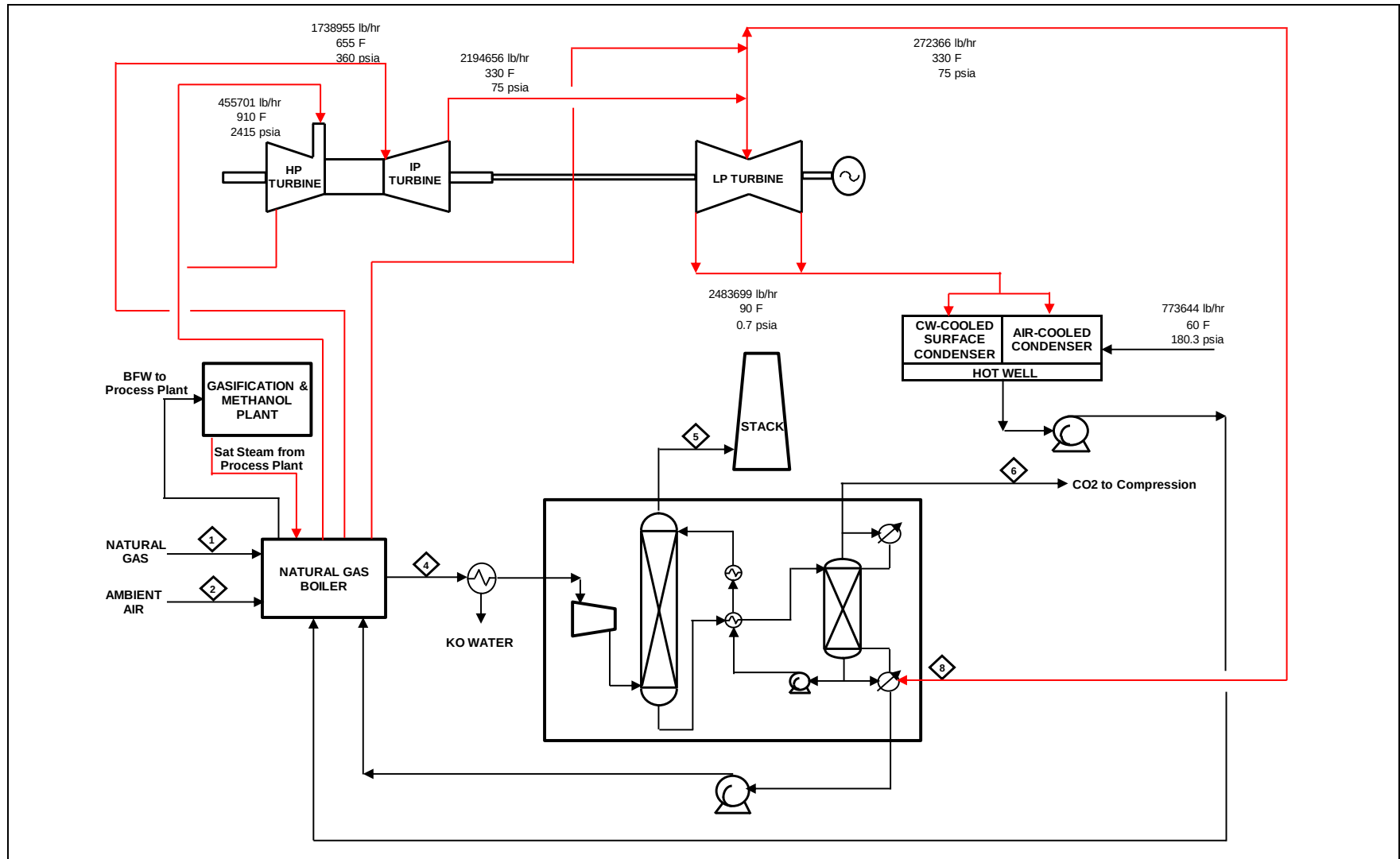


Table 6-8
Case 3d Plant – NGCC Stream Table

STREAM	1	2	4	5	6	7	8
Description	Ambient Air	Natural Gas	Flue Gas to MEA	Treated Gas	CO2 Out	CO2 Product	LP Steam
Mole Flow (Vapor/Liquid)/lbmol/hr							
AR	1057	0	1057	1057	0	0	0
CH4	0	3659	0	0	0	0	0
CO	0	0	0	0	0	0	0
CO2	38	39	4133	413	3719	3719	0
COS	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0
H2O	1128	0	9012	3622	45	0	15119
H2S	0	0	0	0	0	0	0
N2	88394	63	88457	88457	0	0	0
NH3	0	0	0	0	0	0	0
O2	23712	0	15714	15714	0	0	0
SO2	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0
Methanol -- CH3OH	0	0	0	0	0	0	0
Ethanol -- C2H5OH	0	0	0	0	0	0	0
Propanol -- C3H8O	0	0	0	0	0	0	0
Ethane -- C2H6	0	126	0	0	0	0	0
Propane -- C3H8	0	28	0	0	0	0	0
n-Butane -- C4H10	0	16	0	0	0	0	0
Total V/L Flow, lbmol/hr	114330	3930	118374	109265	3764	3719	15119
Total V/L Flow, lb/hr	3299204	68087	3367302	3106521	164499	163687	272366
Total Flow (Solids + V/L), lb/hr	3299204	68087	3367302	3106521	164499	163687	272366
Pressure, psia	15	475	15	15	24	2215	260
Temperature, F	59	60	277	85	69	124	35

6.3 SPARING PHILOSOPHY

The sparing philosophy for the RTI-provided equipment for the Case 3d plant is provided below. Single train designs were utilized throughout with exceptions where equipment has exceeded its maximum capacity limit or dimensions for acceptable transportation to the site and thus required an additional train. The design has:

- One WDP train (1 x 100%)
- One DSRP train (1 x 100%)
- Two AFWGS trains (2 x 50%)
- Two AACRP trains (2 x 50%)

6.4 PERFORMANCE RESULTS

The Nexant-modeled Case 3d plant consumes 19,418 tpd of PRB coal and 37.0 MMSCFD of natural gas at the Midwestern site and produces 927,811 lb/hr (10,100 mtpd) of crude methanol on a net basis. Overall performance for the Case 3d plant is summarized in Table 6-9 which includes the plant's auxiliary power requirements. Loads that are unique to the RTI WDP system are shown in bold and italicized.

Table 6-9
Case 3d Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case 3d
Gas Turbine Power	127,087
Steam Turbine Power	292,806
TOTAL POWER, kWe	419,893
CTM Plant Auxiliary Load Summary, kWe	
Coal Handling	9,090
Slag Handling	2,377
Incinerator Air Blower	5,066
Incinerator Recycle Blower	1,170
Air Separation Unit	174,802
Quench Water Pump	1,830
Syngas Recycle Compressor	7,650
RTI WDP	9,267
RTI DSRP	724
AACRP	29,144
AACRP and NGCC CO ₂ Compressor Auxiliaries	75,611
Methanol Synthesis Syngas Compressor	18,552
Recycle Gas Compressor	2,977
Air Cooler Fans	1,648
Water Treatment	3,124
Miscellaneous BOP	5,000
Circulating Water Pumps	8,445
Ground Water Pumps	895
Cooling Tower Fans	5,506
SUBTOTAL CTM PLANT AUXILIARIES, kWe	362,911
NGCC Plant Auxiliary Load Summary, kWe	
Gas Turbine Auxiliaries	782
Steam Turbine Auxiliaries	106
Transformer Losses	2,303
Miscellaneous BOP	500
Air Cooled Condenser Fans	5,244
Condensate Pumps	845
Boiler Feed Water Pumps	18,554
Amine System Auxiliaries	3,893
SCR	16
SUBTOTAL POWER PLANT AUXILIARIES, kWe	32,244
TOTAL AUXILIARIES, kWe	395,155
NET PLANT POWER, kWe	24,738

Table 6-10 shows the carbon balance for the Case 3d plant. Carbon leaves the CTM process plant as unburned carbon in the slag, CO₂ in the dryer exhaust gas, ASU vent gas, crude methanol product, and the AACRP CO₂ capture product. Additionally, carbon also leaves the power cycle as CO₂ in the stack and Econamine CO₂ capture product. Carbon in the crude methanol is considered to be product and not emissions. The carbon capture efficiency is defined as the amount of carbon in the product streams, which include the crude methanol, the dried and compressed CO₂ products from the AACRP and Econamine CO₂ capture processes and gypsum, relative to the amount of carbon in the coal and natural gas feedstock less carbon contained in the slag. For Case 3d, the carbon capture efficiency is 93.4% vs 92.1 for Case 3a.

Table 6-10
Case 3d Plant – Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	810,195	
ASU Air	598	
Coal Dryer Air	72	
Coal Dryer Supplemental Natural Gas	1,627	
Power Cycle Natural Gas	49,186	
Power Cycle Combustion Air	450	
DSRP Air	19	
ASU Vent		598
Coal Dryer Exhaust Gas		50,099
Carbon in Slag		4,051
Sulfur Product		0
DSRP Vent		3,033
Gypsum		982
Crude Methanol Product		342,955
AACRP CO ₂ Product		410,660
Power Cycle CO ₂ Product		44,672
Power Cycle Exhaust Gas		4,964
CO ₂ in LTGC Condensate (Vented)		141
Convergence Tolerance		-8
Total	862, 148	862,148

Table 6-11 shows the sulfur balance for the Case 3d plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the elemental sulfur recovered in the DSRP, sulfur recovered in the form of gypsum leaving the lime scrubber, sulfur emitted in the coal dryer exhaust gas, and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible. The net sulfur emissions include only the sulfur emitted in the coal dryer exhaust gas. Based on this, the net sulfur emissions for the plant are 0.0001 lb of SO₂/MMBtu. **This level of sulfur emissions not only meets the BACT environmental design basis in Table 2-7, but is over 200 times lower than the sulfur emissions for Case 3a. This demonstrates that RTI's advanced warm syngas cleanup process results in a significant decrease in sulfur emissions. At this low level, no additional sulfur removal would be expected for permitting this plant.**

Table 6-11
Case 3d Plant – Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	11,772	
Sulfur Product		4,413
Gypsum		7,335
Coal Dryer Exhaust Gas		1
CO ₂ Product		22
Convergence Tolerance		1
Total	11,772	11,772

Table 6-12 shows the overall water balance for the Case 3d plant. Raw water is obtained from groundwater (50%) and from municipal sources (50%). Some water is discharged by the processes as effluent suitable for internal consumption. This effluent is internally recycled and consumed by the slag handling process, as wash water to the AACRP, and as makeup to the cooling tower. Some water is discharged from the process to a permitted outfall. Waste water is discharged by the slag handling process and as cooling tower blowdown.

Table 6-12
Case 3b Plant – Overall Water Balance

Water Use, gpm	Raw Water Withdrawal	Process Effluent Produced for Internal Consumption	Internal Consumption	Process Water Discharge
Slag Handling	0	0	353	(353)
Quench Cooler	2,438	0	0	0
DSRP Air Compressor Knockout	0	(1)	0	0
Syngas Cooling Knockout	0	(2,900)	0	0
AACRP Wash Water	0	0	151	0
CO ₂ Compression Knockout	0	(140)	0	0
Steam Cycle Makeup	1,546	0	0	0
Steam Cycle Blowdown	0	(36)	0	0
Flue Gas Cooling Knockout	0	(192)	0	0
Cooling Tower Makeup	5,846	0	2,765	0
Cooling Tower Blowdown	0	0	0	(1,952)
Total	9,830	(3,270)	3,270	(2,305)

Positive values represent consumption while negative values represent production

6.5 EQUIPMENT LIST

The list in Table 6-13 below shows the Case 3d RTI-related equipment that differs from the reference Case 3a reference plant.

Table 6-13
Case 3d Equipment List

ACCOUNT 5A GAS CLEANUP & PIPING
Subaccount 5A.1 AACRP CO₂ Removal System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	Rich Amine Absorber	Packed Column	247,000 cuft/hr syngas	6	0
2	Lean Amine Absorber	Packed Column w/ Wash Trays	637,200 cuft/hr syngas	2	0
3	Amine Stripper	Packed Column	1,320,100 cuft/hr gas flow	2	0
4	LP Flash	Packed Column	2,232,000 cuft/hr gas flow	6	0
5	LP Flash Reflux Drum	Reflux Drum	2,185,000 cuft/hr CO ₂	6	0
6	Lean/Rich Amine Exchanger	Shell and Tube	111.2 MMBtu/hr	2	0
7	Amine Stripper Reboiler	Kettle Reboiler	41.4 MMBtu/hr	2	0
8	Lean Amine Cooler	Shell and Tube	64.3 MMBtu/hr	2	0
9	Flash Overhead Condenser	Shell and Tube	16.9 MMBtu/hr	6	0
10	Lean Amine Pump	Centrifugal	1,260 gpm	6	0
11	Semi-lean Amine Pump	Centrifugal	11,200 gpm	6	0
12	Rich Amine Pump	Centrifugal	3,740 gpm	2	0
13	Flash Reflux Pump	Centrifugal	29 gpm	6	0
14	Amine Stripper Bottoms Pump	Centrifugal	3,610 gpm	2	1
15	Filter Pump	Centrifugal	401 gpm	2	0
16	MP Flash Drum	Reflux Drum	203,700 cuft/hr	2	0
17	Recycle Gas Compressor	Centrifugal	122 acfm @ 456 psia	2	0
18	Amine Tank	Vertical Vessel	10,000 gallons	2	0
19	Filter Package	Filter Package	401 gpm	2	0
18	Filter Package Vessel	Vessel	2,000 gallons	3	0

ACCOUNT 5A GAS CLEANUP & PIPING
Subaccount 5A.2 RTI DSRP System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	DSRP Fixed Bed Reactor	Packed Bed Reactor	350,800 lb/hr	1	0
2	Tail Gas Oxidation Reactor	Packed Bed Reactor	394,200 lb/hr	1	0
3	DSRP Feed Gas Preheater	Shell and Tube	22 MMBtu/hr	1	0
4	Sulfur Condenser	Shell and Tube	4,420 lb/hr sulfur	1	0

5	Oxidation Reactor Gas Preheater	Shell and Tube	8.2 MMBtu/hr	1	0
6	Liquid Sulfur Separator	Pressure Vessel	4,420 lb/hr sulfur	1	0
7	Recycle Gas Compressor	Centrifugal	2,000 acfm @ 458 psia	1	0
8	Pulsing Gas Compressor	Centrifugal	Average 24 acfm @ 524 psia	1	0
9	Scrubber Recycle Cooler	Shell and Tube	78 MMBtu/hr	1	0
10	SO ₂ Scrubber	Tray Column	410,100 lb/hr	1	0
11	Recycle Pump	Centrifugal	1,875 gpm @ 166 ft H ₂ O	1	0
12	Lime Slurry Pump	Centrifugal	350 gpm @ 1,100 ft H ₂ O	1	0
13	Gypsum Filter	Bag Filter	2,000 gpm	1	0
14	Lime Makeup Drum	Pressure Vessel	177,300 lb/hr	1	0

ACCOUNT 5A GAS CLEANUP & PIPING
Subaccount 5A.4a RTI WDP System

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	WDP Adsorption Reactor	Proprietary Transport Reactor	66,200 acfm	1	0
2	WDP Regeneration Reactor	Proprietary Transport Reactor	1,780 acfm	1	0
3	Adsorber Cyclone Separator	Cyclone Separator	66,200 acfm	1	0
4	Stripper Cyclone Separator	Cyclone Separator	1,780 acfm	1	0
5	Adsorber Filter	Candle Filter	66,200 acfm	1	0
6	Regenerator Offgas Filter	Candle Filter	1,780 acfm	1	0
7	Adsorber Filter Lock Hopper + Fines Bin		Proprietary	1	0
8	Regenerator Filter Lock Hopper + Fines Bin		Proprietary	1	0
9	Sorbent Feeder Package (incl Hopper)		Proprietary	1	0
10	Regenerator Air Heat Exchanger	Shell and Tube	76,300 lb/hr	1	0
11	Recycle Syngas Compressor	Centrifugal	380 acfm @ 534 psia	1	0
12	Regenerator Air Compressor	Centrifugal	30,700 acfm @ 15 psia	1	0
13	DSRP Offgas Compressor	Centrifugal	190 acfm @ 505 psia	1	0
14	Syngas Recycle Compressor	Centrifugal	290 acfm @ 598 psia	1	0

ACCOUNT 5A SYNGAS CLEANUP**Subaccount 5A.4b Shift Reactors**

Equipment No.	Description	Type	Design Condition	Operating Qty	Spares
1	AFWGS Reactor 1	Packed Bed Reactor	Proprietary	2	0
2	AFWGS Reactor 2	Packed Bed Reactor	Proprietary	2	0

6.6 CAPITAL COST

Table 6-14 shows the cost breakdown of the Case 3d plant, consistent with the Code of Accounts format as expressed in the Crude Methanol Study. The accounts/subaccounts of interest for this study are:

- 2.3 Dry Coal Injection System,
- 4 Gasifier & Accessories,
- 5A Gas Cleanup & Piping and
- 5B CO₂ Removal and Compression

These are shown with more detail to include the various subaccounts and to provide more clarity as to wherein the major cost differences lie among the cases.

Table 6-15 shows the calculation and addition of owner's costs to determine the TOC, which is used to calculate the methanol RSP.

The estimated TOC of the Case 3d plant in 2011 dollars is \$5,548MM.

Table 6-14
Case 3d Total Plant Cost Summary

Case 3d: Shell Gasifier with RTI WDP CTM Plant Low Rank Western Coal Baseline Study			PRB	Coal Feed, lb/hr Coal HHV, Btu/lb		1,618,190 8,564	Plant Size			10100 metric tons per day		
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost \$	Eng'g CM H.O & Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect				Process	Project	\$	
	1 COAL & SORBENT HANDLING	\$36,523	\$6,549	\$28,311	\$0	\$0	\$71,383	\$6,479	\$0	\$15,572	\$93,434	
	2 COAL & SORBENT PREP & FEED											
	2.3 Dry Coal Injection System	\$203,470	\$0	\$137,594	\$0	\$0	\$341,064	\$29,375	\$0	\$74,088	\$444,528	
	2.x Other Coal & Sorbent Prep & Feed Systems	\$151,813	\$28,795	\$43,835	\$0	\$0	\$224,443	\$19,633	\$0	\$48,816	\$292,892	
	SUBTOTAL 2.	\$355,283	\$28,795	\$181,429	\$0	\$0	\$565,507	\$49,008	\$0	\$122,904	\$737,420	
	3 FEEDWATER & MISC BOP SYSTEMS	\$30,144	\$8,619	\$18,499	\$0	\$0	\$57,262	\$5,378	\$0	\$14,233	\$76,873	
	4 GASIFIER & ACCESSORIES											
	4.1 Gasifier, Syngas Cooler & Auxiliaries (Shell)	\$429,138	\$0	\$194,177	\$0	\$0	\$623,315	\$55,887	\$89,766	\$118,739	\$887,707	
	4.3 ASU/Oxidant Compression	\$600,012	\$0	\$0	\$0	\$0	\$600,012	\$58,159	\$0	\$65,817	\$723,988	
	4.4 LT Heat Recovery & FG Saturation	\$83,110	\$0	\$26,384	\$0	\$0	\$109,494	\$10,686	\$0	\$24,036	\$144,216	
	4.6 Flare Stack System	\$0	\$1,462	\$666	\$0	\$0	\$2,128	\$204	\$0	\$466	\$2,798	
	4.9 Gasification Foundations	\$0	\$31,373	\$18,710	\$0	\$0	\$50,083	\$4,585	\$0	\$13,667	\$68,335	
	SUBTOTAL 4.	\$1,112,260	\$32,835	\$239,937	\$0	\$0	\$1,385,032	\$129,521	\$89,766	\$222,724	\$1,827,043	
	5A GAS CLEANUP & PIPING											
	5A.1 AACRP	\$44,777	\$0	\$37,673	\$0	\$0	\$82,450	\$7,789	\$16,490	\$21,346	\$128,075	
	5A.2 RTI DSRP System	\$12,974	\$0	\$14,363	\$0	\$0	\$27,337	\$2,655	\$5,467	\$7,092	\$42,552	
	5A.3 Mercury Removal	\$7,601	\$0	\$5,745	\$0	\$0	\$13,346	\$1,289	\$667	\$3,060	\$18,362	
	5A.4a RTI WDP System	\$45,570	\$0	\$54,684	\$0	\$0	\$100,253	\$9,748	\$20,051	\$26,010	\$156,062	
	5A.4b Shift Reactors	\$6,501	\$0	\$2,600	\$0	\$0	\$9,101	\$872	\$0	\$1,995	\$11,968	
	5A.5 Syngas Compressor	\$5,645	\$0	\$4,273	\$0	\$0	\$9,918	\$944	\$0	\$1,087	\$11,950	
	5A.6 Blowback Gas Systems	\$6,751	\$1,136	\$640	\$0	\$0	\$8,527	\$809	\$0	\$1,867	\$11,203	
	5A.7 Fuel Gas Piping	\$0	\$2,897	\$1,895	\$0	\$0	\$4,792	\$444	\$0	\$1,047	\$6,282	
	5A.9 HGCU Foundations	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
	SUBTOTAL 5.	\$129,819	\$4,033	\$121,873	\$0	\$0	\$255,724	\$24,551	\$42,675	\$63,504	\$386,455	
	5B CO2 REMOVAL & COMPRESSION											
	5B.1 NGCC CO2 Removal System	\$22,922	\$0	\$6,907	\$0	\$0	\$29,829	\$2,489	\$5,966	\$7,658	\$45,942	
	5B.2 CO2 Compression & Drying (AACRP & NGCC)	\$73,882	\$0	\$27,853	\$0	\$0	\$101,735	\$9,747	\$0	\$22,296	\$133,778	
	5B.3 CO2 Compression & Drying (NGCC)	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
	SUBTOTAL 5B.	\$96,804	\$0	\$34,760	\$0	\$0	\$131,565	\$12,236	\$5,966	\$29,954	\$179,720	
	5C METHANOL PRODUCTION	\$131,805	\$55,452	\$110,906	\$0	\$0	\$298,163	\$29,817	\$0	\$65,596	\$393,577	
	6 COMBUSTION TURBINE/ACCESSORIES	\$47,397	\$394	\$3,814	\$0	\$0	\$51,605	\$11,171	\$11,909	\$14,852	\$89,537	
	7 HRSG, DUCTING & STACK	\$23,807	\$665	\$5,655	\$0	\$0	\$30,128	\$2,501	\$0	\$3,281	\$35,910	
	8 STEAM TURBINE GENERATOR	\$116,640	\$1,181	\$29,739	\$0	\$0	\$147,561	\$13,220	\$0	\$31,866	\$192,646	
	9 COOLING WATER SYSTEM	\$18,894	\$20,659	\$15,301	\$0	\$0	\$54,854	\$5,074	\$0	\$12,174	\$72,102	
	10 ASH/SPENT SORBENT HANDLING SYS	\$82,188	\$3,649	\$77,402	\$0	\$0	\$163,239	\$15,794	\$0	\$18,685	\$197,719	
	11 ACCESSORY ELECTRIC PLANT	\$40,384	\$23,248	\$40,808	\$0	\$0	\$104,440	\$9,072	\$0	\$22,589	\$136,101	
	12 INSTRUMENTATION & CONTROL	\$12,576	\$3,801	\$10,547	\$0	\$0	\$26,925	\$2,417	\$1,346	\$5,499	\$36,187	
	13 IMPROVEMENTS TO SITE	\$5,862	\$2,880	\$15,342	\$0	\$0	\$24,084	\$2,378	\$0	\$7,939	\$34,401	
	14 BUILDINGS & STRUCTURES	\$0	\$7,898	\$10,942	\$0	\$0	\$18,840	\$1,703	\$0	\$3,332	\$23,874	
CALCULATED TOTAL COST		\$2,240,387	\$200,659	\$945,265	\$0	\$0	\$3,386,312	\$320,320	\$151,662	\$654,704	\$4,512,998	

Table 6-15
Case 3d Total Plant Cost Summary

Owner's Costs	\$ x \$1,000
Preproduction Costs	
6 months All Labor	\$31,701
1 Month Maintenance Materials	\$6,040
1 Month Non-Fuel Consumables	\$1,971
1 Month Waste Disposal	\$1,259
25% of 1 Months Fuel Cost at 100% CF	\$7,178
2% of TPC	\$90,260
Total	\$138,409
Inventory Capital	
60 day supply of fuel at 100% CF	\$56,640
60 day supply of non-fuel consumables at 100% CF	\$2,446
0.5% of TPC (spare parts)	\$22,565
Total	\$81,651
Initial Cost for Catalyst and Chemicals	\$14,989
Land	\$900
Other Owner's Cost	\$676,950
Financing Costs	\$121,851
Total Owner's Costs	\$1,034,750
Total Overnight Costs (TOC)	\$5,547,748

6.7 OPERATING COSTS

Table 6-16 shows the OPEX breakdown for the Case 3d Plant.

Table 6-16
Case 3d Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Case 3d - Shell Gasifier with RTI WDP CTM Plant				
Plant Size (mtpd Methanol)	10,100				
Primary/Secondary Fuel:	PRB/Natural Gas				
Design/Construction	5 years		Book Life (yrs):	20	
TPC (Plant Cost) Year	June 2011		TPI Year:	2016	
Capacity Factor (%)	90		CO2 Captured (TPD)	18053	
OPERATING & MAINTENANCE LABOR					
Operating Labor					
Operating Labor Rate (base):	\$39.70 \$/hr				
Operating Labor Burden:	30.00 % of base				
Labor Overhead Charge	25.00 % of labor				
Operating Labor Requirements per Shift	units/mod		Total Plant		
Skilled Operator	2.0		2.0		
Operator	10.0		10.0		
Foreman	1.0		1.0		
Lab Tech's etc	3.0		3.0		
TOTAL Operating Jobs	16.0		16.0		
Annual Cost					
\$					
Annual Operating Labor Cost	\$7,233,658				
Maintenance Labor Cost	\$43,487,251				
Administration & Support Labor	\$12,680,227				
Property Taxes and Insurance	\$90,259,964				
TOTAL FIXED OPERATING COSTS	\$153,661,100				
VARIABLE OPERATING COSTS					
Maintenance Material Cost					
					\$65,230,876
Consumables	Consumption		Unit	Initial Fill	
	Initial	/Day	Cost	Cost	
Water(/1000 gallons)	0	7,078	1.67	\$0	\$3,891,826
Chemicals					
MU & WT Chem (lb)	0	42171	0.27	\$0	\$3,710,475
Carbon (Hg Removal) (lb)	263903	402	1.63	\$430,162	\$215,253
MEA Solvent (ton)	152	0.21	3751.70	\$569,647	\$255,903
Water Gas Shift Catalyst (lb)	499624	273.77	5.64	\$2,818,112	\$507,260
WDP Sorbent (lb)	Proprietary	Proprietary	Proprietary	\$7,211,168	\$6,399,977
AACRP Solvent	382034	209.33	2.80	\$1,068,882	\$192,399
Methanol Synthesis Catalyst	4014	3.66	534.68	\$2,146,441	\$643,218
DSRP Catalyst	57860	31.70	11.15	\$645,139	\$116,125
Oxidation/Reduction Catalyst	63800	34.96	1.56	\$99,528	\$17,915
Calcium Hydroxide	5200	425400	0.04	\$199	\$5,339,775
Subtotal Chemicals				\$14,989,278	\$17,398,299
Other					
Supplemental Electricity (MWh consumed)	0	0	62.33	\$0	\$0
Gases, N2 etc.(/100scf)	0	0	0.00	\$0	\$0
LP Steam (/1000 lbs)	0	0	0.00	\$0	\$0
Subtotal Other				\$0	\$0
Waste Disposal:					
Spent Mercury Catalyst (lb)	0	402	0.65	\$0	\$85,837
Flyash (ton)	0	0	0.00	\$0	\$0
Slag (ton)	0	1638	25.11	\$0	\$13,515,321
Subtotal Waste Disposal				\$0	\$13,601,158
By-products & Emissions					
Sulfur (tons)	0	0.0	0.00	\$0	\$0
Supplemental Electricity (MWh generated)	0	594	-59.59	\$0	-\$11,622,031
Subtotal By-Products				\$0	-\$11,622,031
TOTAL VARIABLE OPERATING COSTS				\$14,989,278	\$88,500,129
Coal (tons)	0	19,418	36.57	\$0	\$233,276,555
Natural Gas (MMBtu)	0	38,152	6.13	\$0	\$76,827,051

6.8 METHANOL PRODUCT REQUIRED SELLING PRICE

Table 6-17 shows a summary of the power output, CAPEX, OPEX, and methanol product RSP for the Case 3d Plant. The Case 3d Plant methanol RSP is estimated to be \$1.41/gal under the loan guarantee finance structure and \$1.66/gal under the commercial fuels finance structure.

Table 6-17
Plant Performance and Economic Summary

Case	Case 3d	
CAPEX, \$MM		
Total Installed Cost (TIC)	\$3,386	
Total Plant Cost (TPC)	\$4,513	
Total Overnight Cost (TOC)	\$5,548	
OPEX, \$MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC _{fix})	\$153.7	
Variable Operating Cost, less Fuel (OC _{var})	\$111.2	
Coal Feedstock (OC _{coal})	\$259.2	
Natural Gas Feedstock (OC _{NG})	\$85.4	
Import/(Export) Power (OC _{Power})	(\$12.9)	
Total OPEX	\$596.6	
Plant Output		
Crude Methanol Product, tons per year	4,063,813	
Net Power Output, MWe	24.77	
Required Selling Price^A		
Excluding CO₂ TS&M^{BE}, \$/ton	408.9	481.7
Including CO₂ TS&M^{BE}, \$/ton	426.8	499.6
RSP Component Details (\$/gal)		
Capital ^B	0.85	1.10
Fixed O&M	0.14	
Variable O&M	0.09	
Coal	0.21	
Natural Gas	0.07	
Power	(0.01)	
CO ₂ TS&M	0.06	
RSP^B Total (\$/gal)	1.41	1.66
Costs of CO₂ Captured^{B,C} (\$/tonne)	19.0	21.1
Costs of CO₂ Avoided^{B,D} (\$/tonne)	30.0	32.1

^A Capacity factor assumed to be 90 percent

^B Values shown are for two financial structures

The first (lower value) is based on the loan guarantee finance structure

The second (higher value) is based on the commercial fuels finance structure

^C Excludes CO₂ TS&M

^D Includes CO₂ TS&M

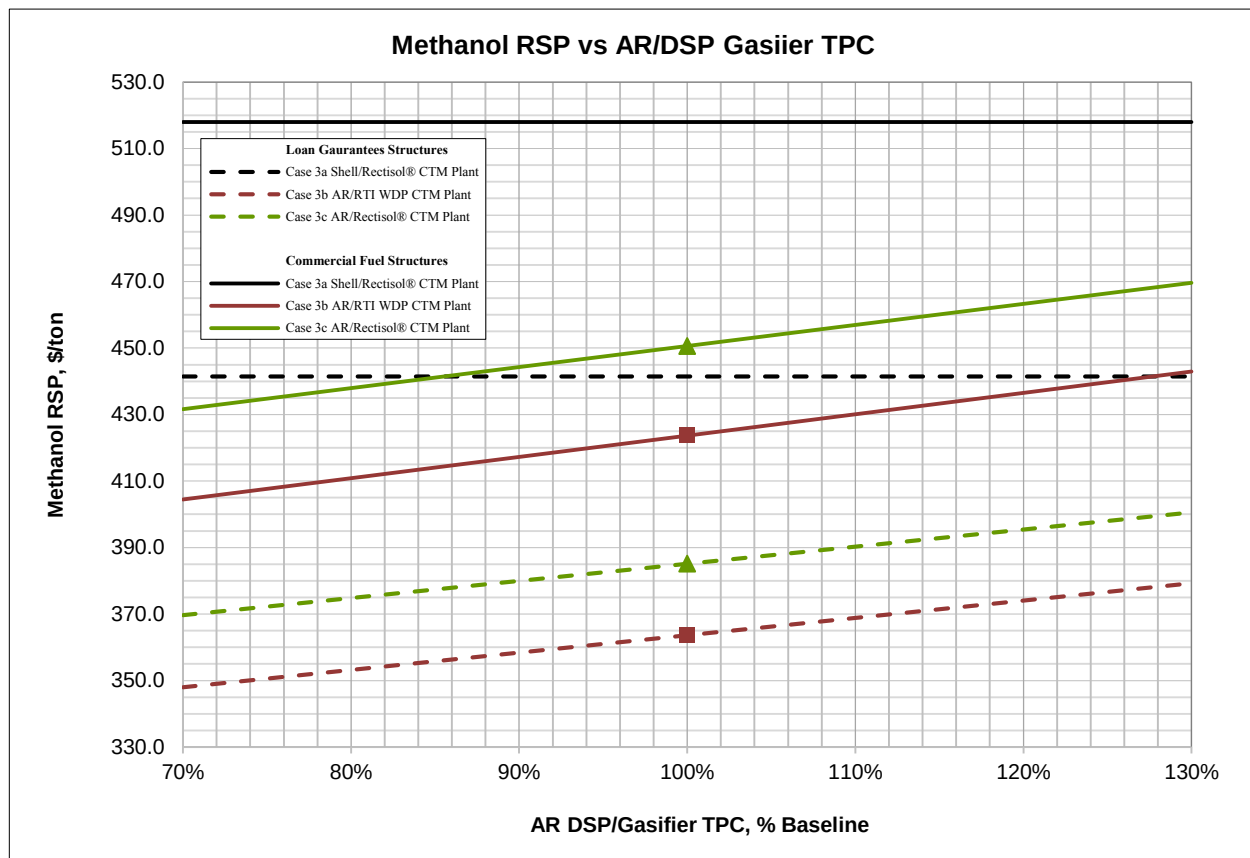
^E Based on 301.73 gallons/ton

Sensitivity analysis was carried out to determine the effects of various parameters of the AR gasifier and RTI WDP systems on the overall methanol product RSP. The parameters investigated include: AR and RTI systems capital cost, CTM plant CF, feedstock (coal and natural gas) cost, electric selling price, CO₂ sales price, cost of CO₂ emissions and CCF

7.1 AR DSP/GASIFIER SYSTEM COST

Figure 7-1 shows how the methanol RSPs for the AR gasifier (3b and 3d) cases change as the AR DSP and gasifier TPC varies from -30% to +30%. Also shown in Figure 7-1 are the reference Case 3a methanol RSPs at \$517.9/ton and \$441.5/ton for the commercial fuels and loan guarantee finance structures respectively.

Figure 7-1
Sensitivity Analysis – RSP vs AR DSP/Gasifier TPC



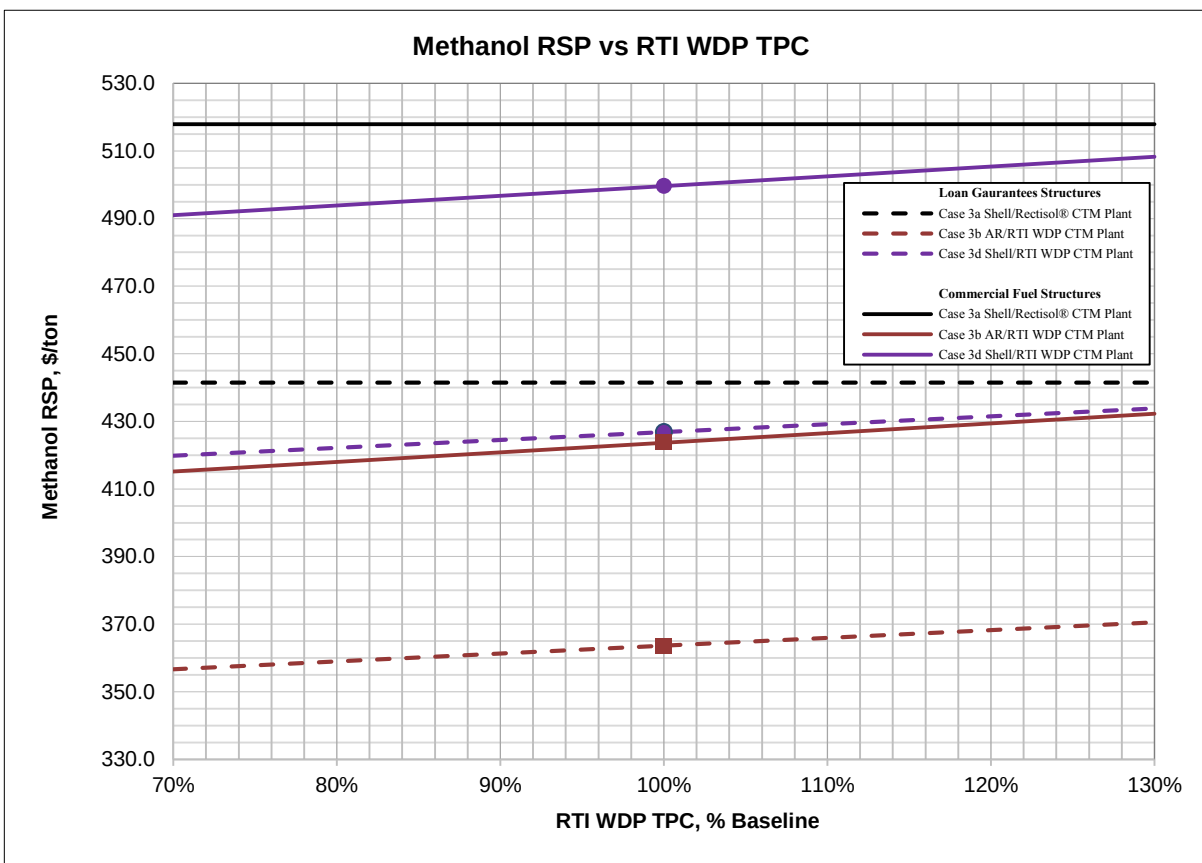
From Figure 7-1, it can be seen that at the high end of the AR DSP and gasifier TPC (+30%), the methanol RSPs for Cases 3b and 3c are still less than the methanol RSP for the Case 3a reference plant.

Under the commercial fuels financing structure, every 1.6% increase in DSP and gasifier TPC, equivalent to about \$12MM, increases the methanol RSP by \$1/ton. Under the loan guarantee financing structure, every 1.9% increase in DSP and gasifier TPC, equivalent to about \$14.5MM, increases the methanol RSP by \$1/ton.

7.2 RTI WDP SYSTEM COST

Figure 7-2 shows how methanol RSPs for the RTI WDP plants (Cases 3b and 3d) change as the RTI WDP system TPC varies from -30% to +30%. The RTI WDP system TPC includes the costs for the RTI WDP, DSRP, AFWGS and AACRP processes. For reference purposes, the Case 3a reference plant methanol RSPs at \$517.9/ton for the commercial fuels and \$441.5/ton for the loan guarantee finance structures are shown in Figure 7-2 as well.

Figure 7-2
Sensitivity Analysis – RSP vs RTI WDP TPC



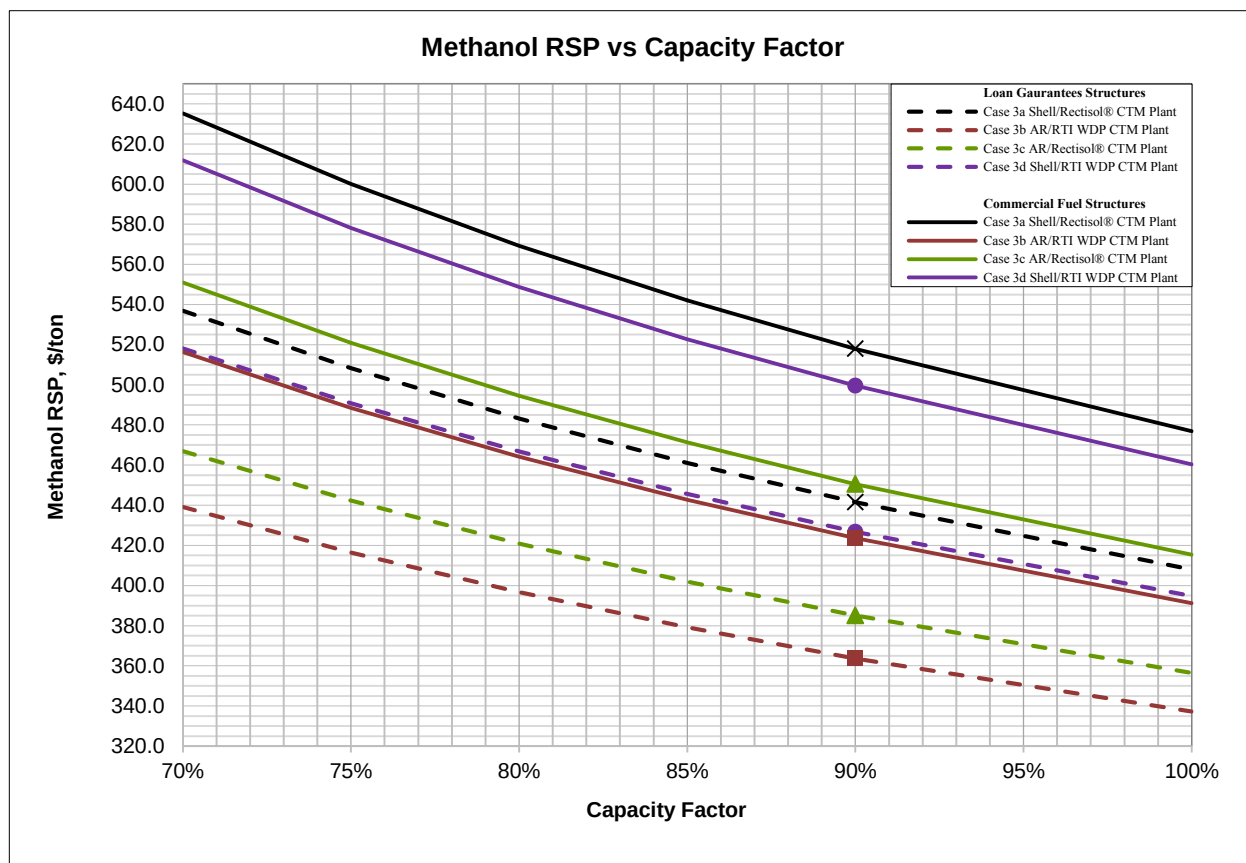
From Figure 7-2, it can be seen that at the high end of the RTI WDP TPC (+30%), the methanol RSPs for Cases 3b and 3d are still less than the RSP for the reference Case 3a reference plant.

Under the commercial fuels financing structure, every 3.5% increase in RTI WDP TPC, equivalent to about \$12MM, increases the methanol RSP by \$1/ton. Under the loan guarantee financing structure, every 4.3% increase in RTI WDP TPC, equivalent to about \$14.5MM, increases the methanol RSP by \$1/ton.

7.3 CAPACITY FACTOR

The baseline CTM plant CF used in this study is 90%. Figure 7-3 shows how the methanol product RSP varies with plant CF as it varies from 70% to 100%.

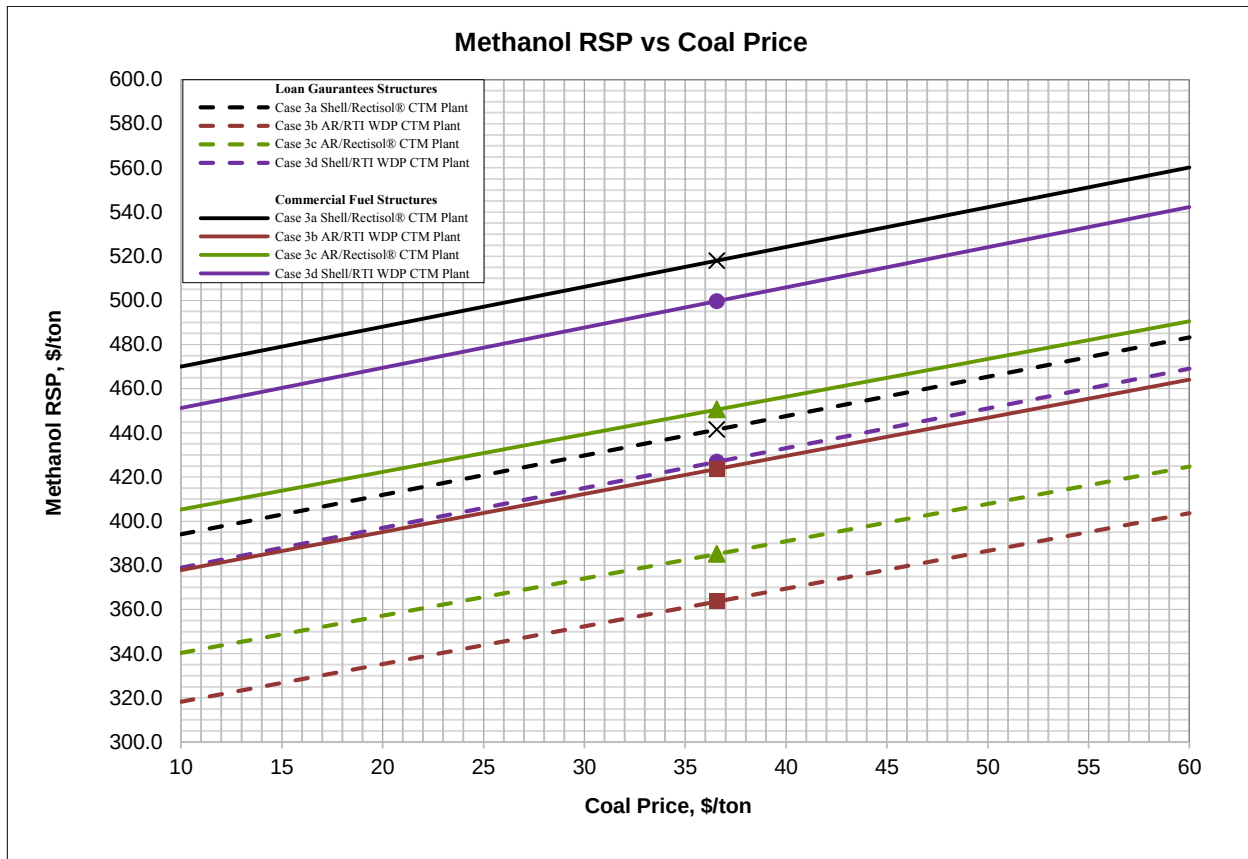
Figure 7-3
Sensitivity Analysis – RSP vs CTM Plant Capacity Factor



7.4 COAL PRICE

The PRB coal feedstock price used in this study was \$36.57/ton, which was specified in the *QGESS Fuel Prices for Selected Feedstocks in NETL Studies* document. Figure 7-4 shows how the methanol RSP varies as the coal price is varied from \$10/ton to \$60/ton.

Figure 7-4
Sensitivity Analysis – RSP vs Coal Feedstock Price



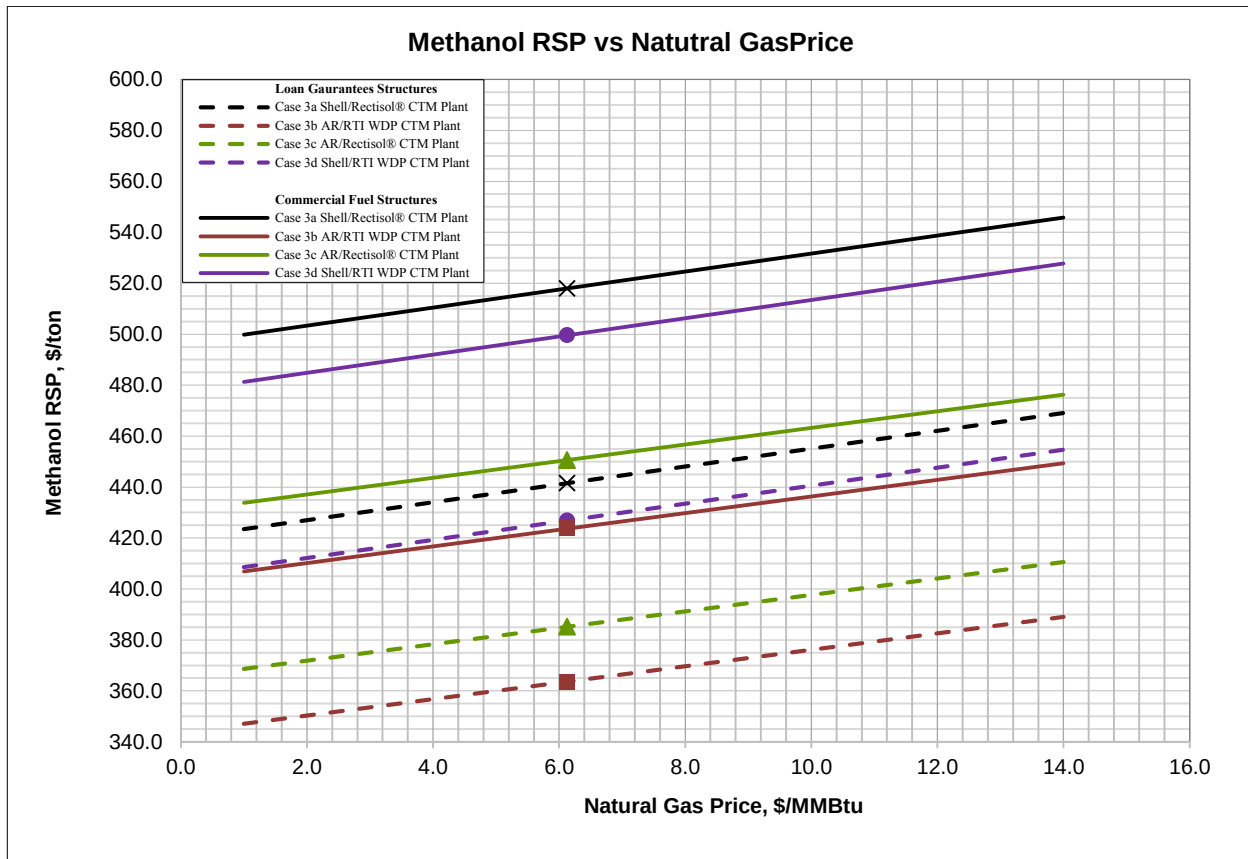
For the Shell gasifier cases (3a and 3d), the methanol RSPs increase at approximately the same rate with increases in coal prices since the feedstock requirements per unit of product are about the same. For these cases, in both commercial fuels and loan guarantee financing structures, the methanol RSP increases by \$1/ton for every \$0.56/ton increase in coal price.

For the AR gasifier cases (3b and 3c), the methanol RSPs also increase at about the same rate with increases in coal prices. However, the RSPs are less sensitive to coal pricing than the Shell gasifier cases since the AR gasifier has a higher cold gas efficiency and uses less coal to produce the same amount of methanol. In both the commercial fuels and loan guarantee financing structures, the methanol RSP for the AR gasifier cases increase by \$1/ton for every \$0.59/ton increase in coal price.

7.5 NATURAL GAS PRICE

Natural gas is used in all cases as fuel to produce power via combined cycle to meet the CTM plant's auxiliary demands. For cases 3a and 3d, it is also used as supplemental fuel to be fired in the coal dryer to dry the as-received coal down to 6% moisture. The natural gas feedstock price used in this study was \$6.13/MMBtu, which was specified in the *QGESS Fuel Prices for Selected Feedstocks in NETL Studies* document. Figure 7-5 shows how the methanol RSP varies with natural gas price as it varies from \$1/MMBtu to \$14/MMBtu.

Figure 7-5
Sensitivity Analysis – RSP vs Natural Gas Feedstock Price



Similar to the coal price sensitivity analysis results, for the Shell gasifier cases (3a and 3d), the methanol RSPs increase at approximately the same rate with increases in natural gas prices since the natural gas requirements per unit of product are about the same. For these cases, in both commercial fuels and loan guarantee financing structures, the methanol RSP increases by \$1/ton for every \$0.28/MMBtu increase in natural gas price.

For the AR gasifier cases (3b and 3c), the methanol RSPs also increase at about the same rate with increases in natural gas prices. However, the RSPs are less sensitive to natural gas pricing than the Shell gasifier cases. The AR cases generate 99.9% of the methanol generated by the Shell cases, but consume about 8% less natural gas than the Shell cases. This reduction in natural gas consumption comes from the fact that the AR gasifier convert more of the coal into methane, which increases the heat content of the purge gas from the methanol process. With a higher heat content, this purge gas eliminates the need for supplemental natural gas for the coal drying process and reduces natural gas consumption in the NGCC.

In both the commercial fuels and loan guarantee financing structures, the methanol RSP for the AR gasifier cases increase by \$1/ton for every \$0.31/MMBtu increase in natural gas price.

7.6 ELECTRICITY PRICE

The combustion of natural gas in the combustion turbine, combined with heat recovery in the HRSG and WHR in the CTM plant from the exothermic methanol synthesis reaction and other processes, result in a substantial amount of excess electricity being generated in Cases 3b, 3c, and 3d. The revenue derived from the sale of this excess electricity is applied to the cash flow and results in a lower RSP. The sensitivity of the RSP to the selling price of electricity is shown in Figure 7-6.

The electricity selling price used in this study was \$59.59/MWh. While the excess power can be sold to the grid, the sale may be at a steep discount as entities that are negotiating a power purchase agreement will know the power production is an inherent by-product of core methanol production operations. Consequently, the actual achieved transfer price for excess power would be highly project dependent. A range of \$0 - \$120/MWh is used in this sensitivity analysis.

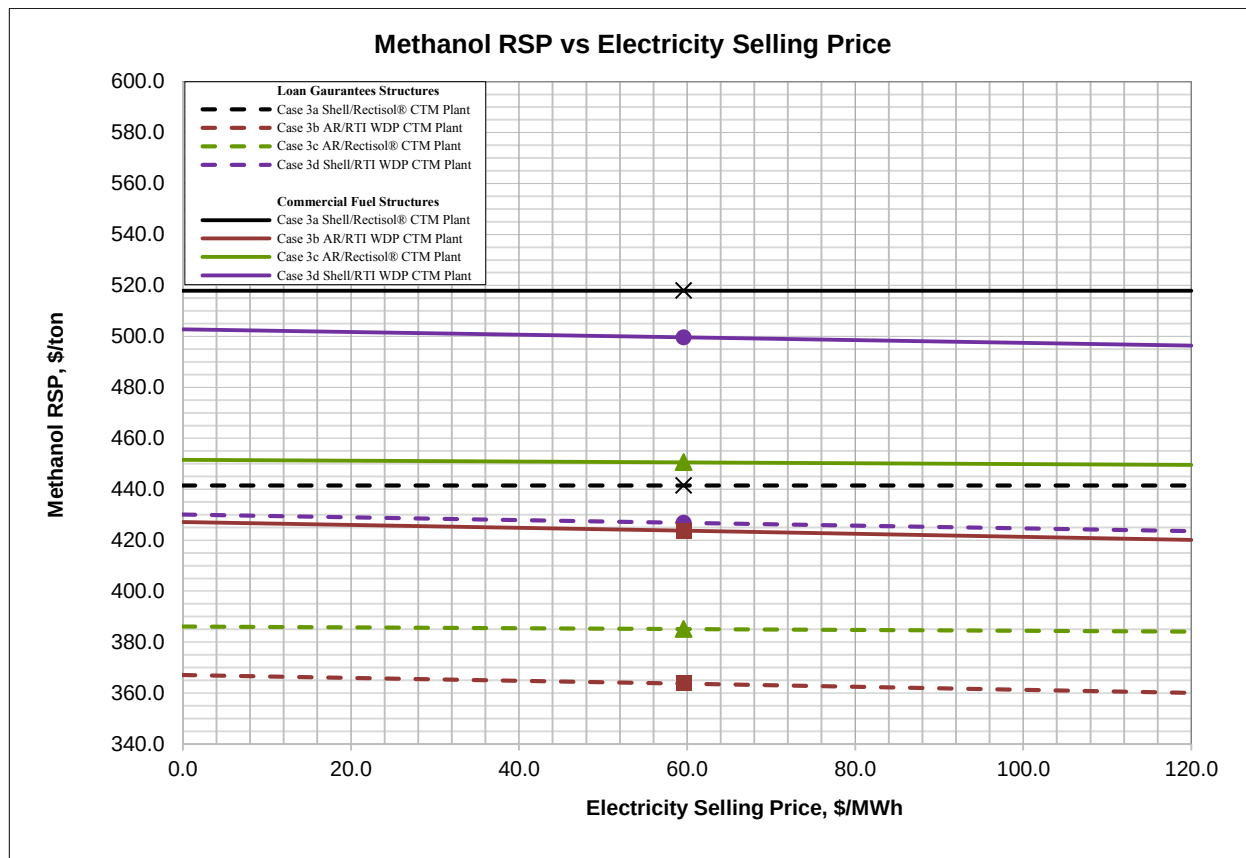
For Case 3a reference plant, there is a net import power of just 0.8 MW. This plant effectively operates as a power-neutral plant and is considered to have no dependency on electricity prices.

The Case 3b plant has the highest net power export (26.9 MW) of the four cases and therefore its methanol RSP has the greatest dependency on electricity selling price. For both the commercial fuels and loan guarantee financing structures, every \$17/MWh increment in electricity selling price decreases the RSP by \$1/ton.

For the Case 3d plant, which net exports 24.7 MW, every \$19/MWh increment in electricity selling price decreases the RSP by \$1/ton. Case 3c, which only exports about 8 MW to the grid after meeting auxiliary plant demand, is much less sensitive to electricity selling price, requiring a \$61/MWh increase to decrease the RSP by \$1/ton.

It should be noted that in the least advantageous scenario where the electricity selling price is \$0/MWh i.e. no revenue is derived from power export to the grid, the methanol RSPs for all three cases are still lower than the reference case. In other words, the CTM plants utilizing the AR gasifier and/or RTI WDP systems still perform better financially than the reference case even when the excess power generated has no value.

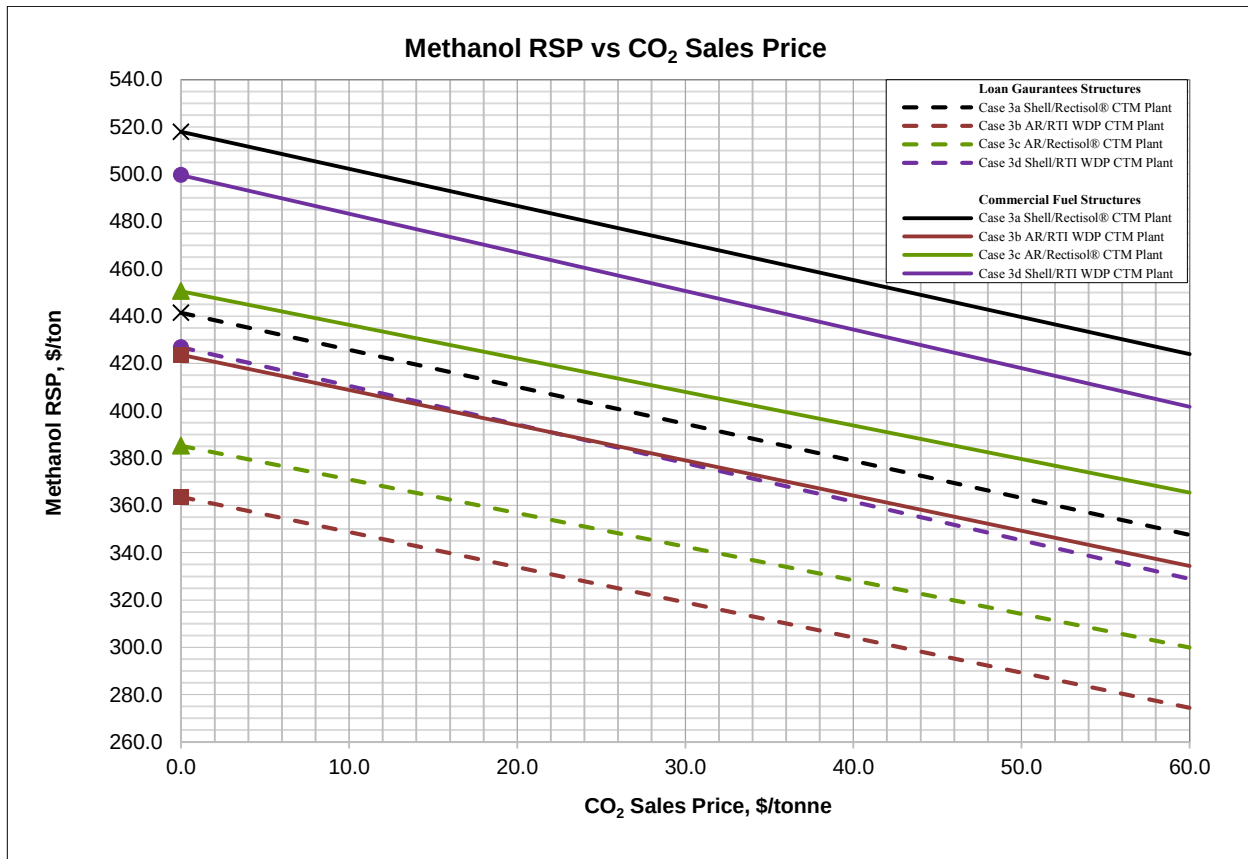
Figure 7-6
Sensitivity Analysis – RSP vs Electricity Price



7.7 CO₂ SALES PRICE

Options for carbon sequestration include both storage in a saline reservoir and usage in enhanced oil recovery (EOR). The impact of selling the captured CO₂ for EOR or other uses at various plant gate sales price, ranging from \$0-\$60/tonne, is shown in Figure 7-6. Sensitivity to CO₂ sales at plant gate prices is shown in Figure 7-6.

Figure 7-7
Sensitivity Analysis – RSP vs CO₂ Sales Price

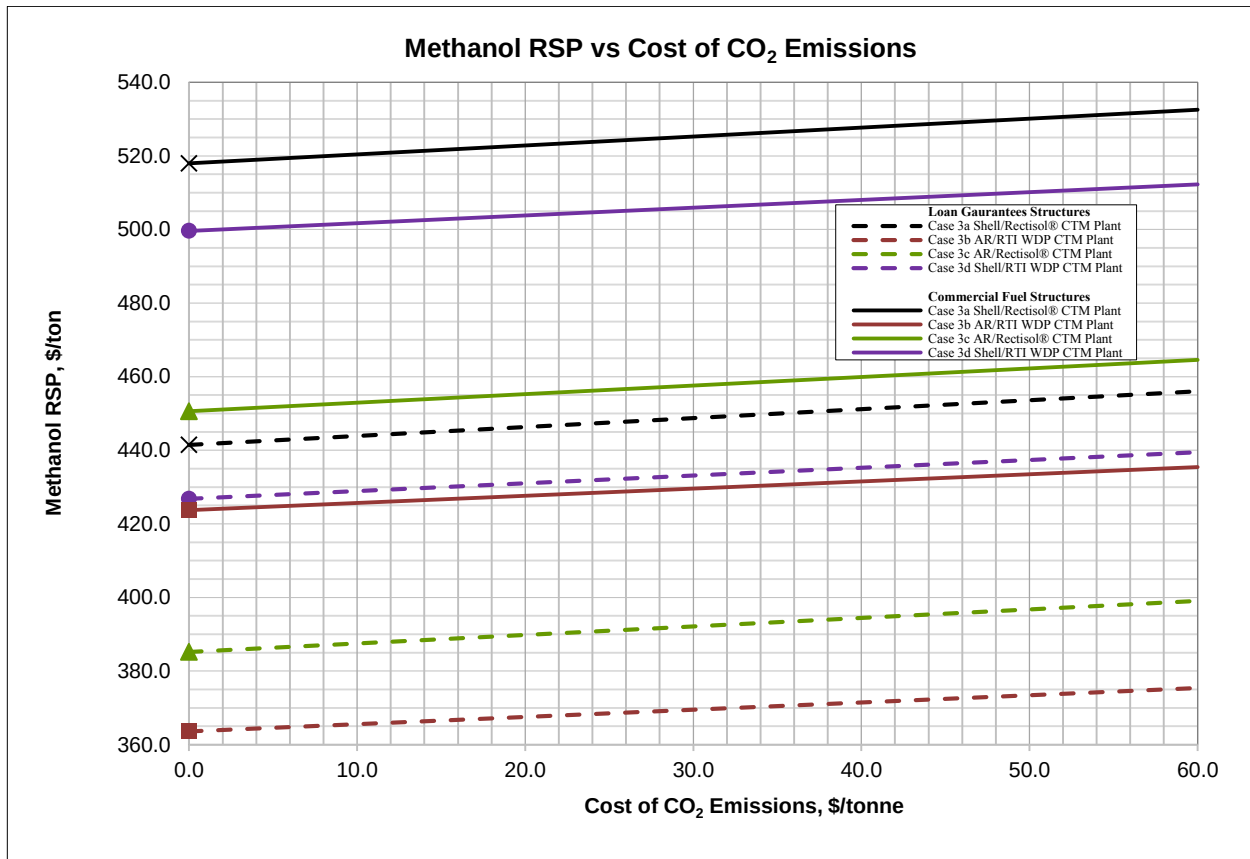


Due to the different rate of CO₂ capture between the Rectisol[®] AGR CTM plants (92.2% capture for Case 3a and 92.1% capture for Case 3c) and RTI WDP (93.4% CO₂ capture for Case 3b and 93.3% CO₂ capture for Case 3d), the slopes are slightly different for the four cases. The cases with more CO₂ capture (Cases 3b and 3d) have a greater decrease in methanol RSP as CO₂ sales price increases.

7.8 COST OF CO₂ EMISSIONS

The sensitivity to CO₂ emissions costs is shown in Figure 7-8. The baseline case assumes that there are no costs associated with venting CO₂ to the atmosphere (\$0/tonne). The cost of CO₂ emissions is subsequently varied to a maximum of \$60/tonne to determine its effect on methanol RSP.

Figure 7-8
Sensitivity Analysis – RSP vs Cost of CO₂ Emissions



For the same explanation provided in Section 7.7 the slopes are slightly different for the four cases due to the different rate of CO₂ capture, hence different CO₂ emissions rate. The cases with more CO₂ capture vent less CO₂ to the atmosphere, so they have a smaller increase in methanol RSP as cost of CO₂ emissions increases.

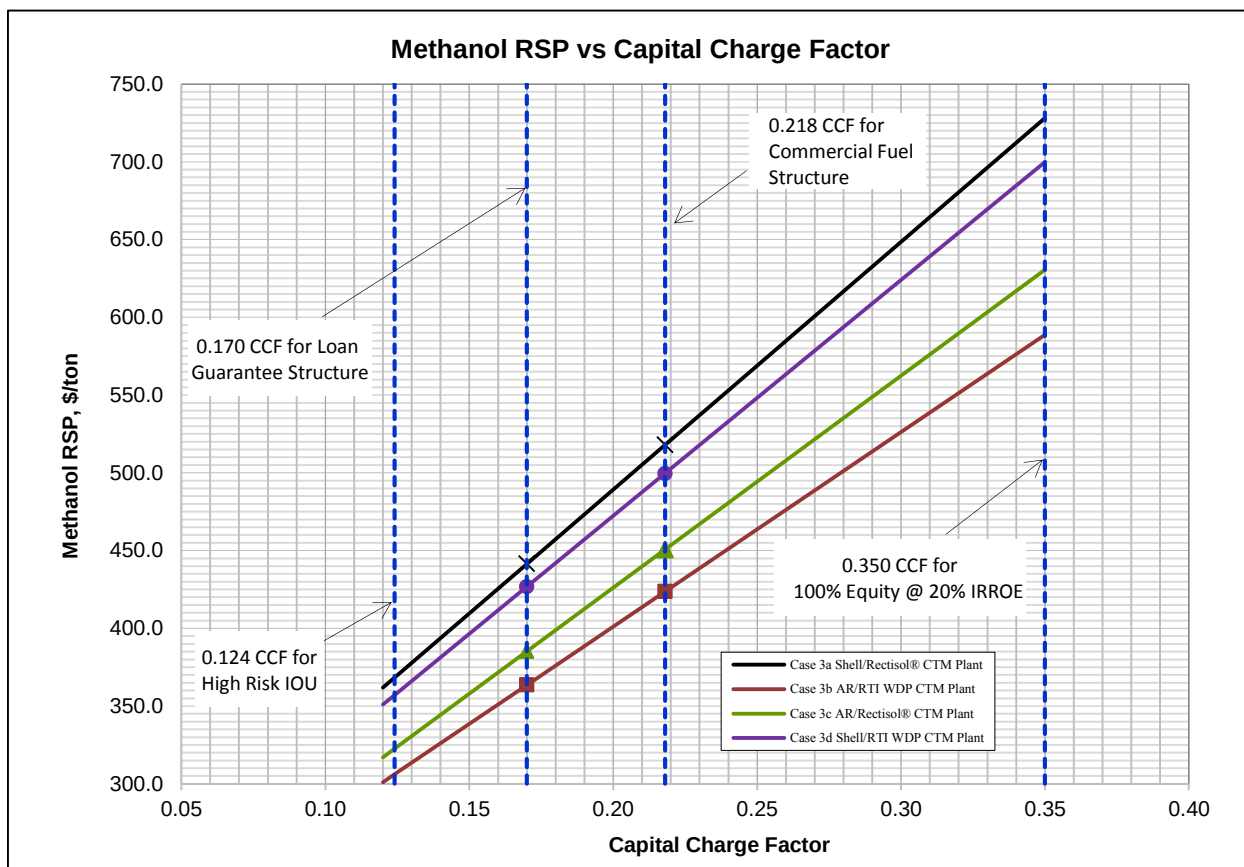
The methanol RSPs are less sensitive to emissions cost than to CO₂ sales price because about 6-8 times more CO₂ is captured by the CTM plant than is vented. The RSP is therefore about 6-8 times more sensitive to CO₂ sales price than to cost of CO₂ emissions.

7.9 CAPITAL CHARGE FACTOR

Two financial structures were assumed for calculating methanol RSPs in this study as described in Section 2.14.1. These structures are based on typical values for fuel projects with and without loan guarantees or government subsidies. The assumed values were used in the PSFM to establish the CCF, which is the portion of the TOC to include in the annual cost of producing a product, for each financial structure. The sensitivity of the methanol RSP to the CCF is illustrated in Figure 7-9. The methanol RSP values were calculated for CCFs ranging from 0.10 to 0.35. The lines denoting CCFs of 0.124 (the value estimated for a high risk investor-owned utility (IOU) project assuming 45 percent debt at 5.5% interest and 55 percent equity and 12%

internal rate of return on equity (IRROE)) and 0.35 (the value estimated for 100 percent equity and 20% IRROE) are also included in the chart.

Figure 7-9
Sensitivity Analysis – RSP vs CCF



7.10 INTEGRATION BENEFITS

Throughout the discussion of the results for each of the sensitivity studies, the primary focus has been sensitivity to the factor being varied. However, another important observation can be made when examining the differences between the cases across the different sensitivity studies. The first pattern that can be seen is that both the AR advanced gasification and RTI advanced syngas cleanup results in reductions in the methanol RSP for every factor tested. Upon closer evaluation, the reduction in methanol RSP is consistently greater when these two advanced technologies are integrated than when they are not integrated. This serves to clearly show the synergistic benefits of integration of these two advanced technologies.

8.1 STUDY OBJECTIVE

The objective of this study was to assess the benefits of integrating the technologies offered by RTI and AR for reducing the capital and production costs for syngas generation and cleanup in a CTM plant with carbon capture.

The reference case for this study was Case 2 from the Crude Methanol Study. The choice of Case 2 from the study was straightforward since it was the only CTM case in this study that included CO₂ capture for sequestration.

To fully evaluate the benefits of integrating these technologies, a total of four case studies were completed. The specific technologies included in each of the four CTM plant configurations are identified in the CTM case study matrix shown in Table 8-1.

Table 8-1
Case Study Matrix for Crude Methanol Production with CO₂ Capture

	Case 3a ¹	Case 3b	Case 3c	Case 3d
Gasification Technology				
Shell Gasification with Lockhopper-Based Feed System	✓			✓
AR Compact Gasifier with Dry Solids Pump (DSP) Feed System		✓	✓	
Gas Cleanup				
Rectisol® for CO ₂ and Sulfur Removal ²	✓		✓	
RTI WDP with AACRP		✓		✓
Water Gas Shift				
Sour Shift	✓		✓	
AFWGS		✓		✓
Methanol Production	✓	✓	✓	✓
NGCC Power Generation with Fluor Econoamine CO ₂ Capture	✓	✓	✓	✓
CO ₂ Drying and Compression (to 2,215 psia)	✓	✓	✓	✓

1 Reference case based on Nexant's benchmark simulation of the Crude Methanol Study Case 2

8.2 RESULTS SUMMARY

Table 8-2 shows a summary comparison of the CAPEX, OPEX, methanol production, plant power import/export, auxiliary power consumption, RSP and cost of CO₂ captured/avoided for the four cases.

Table 8-2
CTM Results Summary

Case	Case 3a	Case 3b	Case 3c	Case 3d
CTM Configuration				
Gasifier	Shell	AR	AR	Shell
Sulfur and CO ₂ Removal	Rectisol®	RTI WDP	Rectisol®	RTI WDP
CAPEX, \$MM				
Total Installed Cost (TIC)	\$3,581	\$2,732	\$2,897	\$3,386
Total Plant Cost (TPC)	\$4,802	\$3,707	\$3,950	\$4,513
Total Overnight Cost (TOC)	\$5,892	\$4,567	\$4,857	\$5,548
OPEX, \$MM/yr (100% CF Basis)				
Fixed Operating Cost (OC _{fix})	\$162.9	\$128.0	\$135.6	\$153.7
Variable Operating Cost, less Fuel (OC _{var})	\$103.9	\$95.3	\$87.9	\$111.2
Coal Feedstock (OC _{Coal})	\$259.2	\$244.8	\$244.8	\$259.2
Natural Gas Feedstock (OC _{NG})	\$85.4	\$77.6	\$78.4	\$85.4
Power Import/(Export)	\$0.4	(\$14.1)	(\$4.0)	(\$12.9)
Total OPEX	\$611.8	\$531.4	\$542.7	\$596.6
Methanol Production, mtpd (100% CF Basis)				
Gross Methanol Production	10,218	10,079	10,204	10,100
Rectisol® Methanol Makeup	8	0	6	0
Net Methanol Production	10,210	10,079	10,198	10,100
NGCC Plant Power Generation, MWe				
Gas Turbine	127.1	127.9	127.8	127.1
Steam Turbine	264.7	239.2	199.1	292.8
Total Power Generation	391.8	367.1	326.8	419.9
Auxiliary Consumption, MWe				
<i>Process Plant</i>				
Coal Handling and Drying	17.4	16.4	16.4	17.7
Gasification and ASU	185.3	163.6	165.1	184.3
Process AGR	42.5	42.6	31.7	39.1
Process CO ₂ Compression	65.2	64.8	52.4	75.6
Methanol Synthesis	27.3	4.7	5.0	23.2
Process Plant BOP	22.5	20.5	20.2	23.0
<i>Total Process Plant Auxiliary Consumption</i>	<i>360.2</i>	<i>312.5</i>	<i>290.9</i>	<i>362.9</i>
<i>NGCC Plant</i>				
NGCC Auxiliary Consumption	22.4	23.5	17.7	28.4
NGCC PCC	3.9	4.1	4.1	3.9
NGCC PCC CO ₂ Compression	6.0	w/Process	6.4	w/Process
<i>Total NGCC Auxiliary Consumption</i>	<i>32.3</i>	<i>27.6</i>	<i>28.2</i>	<i>32.2</i>
Total Auxiliary Consumption	392.6	340.1	319.1	395.1
Net Power Generation/(Consumption)	(0.8)	26.9	7.7	24.8
Fuel Rate and Efficiency				
Coal Feed Rate, tpd AR Coal	19,418	18,339	18,339	19,418
Natural Gas Feed Rate, MMBtu/day	38,152	34,675	35,051	38,152
Total Thermal Input In (LHV), MMBtu/day	354,934	333,984	334,323	354,934
Methanol Fuel Thermal Output (LHV), MMBtu/day	188,547	185,730	188,421	185,877
Power Output, MMBtu/day	(63)	2,207	634	2,029
Thermal Efficiency (LHV), %	53.1%	56.3%	56.5%	52.9%
CO₂ Capture				
CO ₂ Capture Product, tpd	19,413	18,201	17,576	20,021
CO ₂ Vented, tpd	2,966	2,359	2,833	2,538
Carbon Emitted as CO ₂ , % Carbon in Fuel	7.9%	6.6%	7.9%	6.7%

Table 8-2 (cont'd)
CTM Results Summary

Case	Case 3a		Case 3b		Case 3c		Case 3d	
Methanol RSP								
Methanol RSP, excl CO₂ TS&M, \$/ton^{BE}	424.2	500.7	347.3	407.3	359.5	422.7	408.9	481.7
, incl CO₂ TS&M, \$/ton^{BE}	441.5	517.9	363.6	423.7	375.1	438.3	426.8	499.6
RSP Components Details, \$/gal								
Capital ^B	0.90	1.15	0.70	0.91	0.74	0.95	0.85	1.10
Fixed O&M	0.15		0.12		0.12		0.14	
Variable O&M	0.08		0.08		0.07		0.09	
Coal	0.21		0.20		0.20		0.21	
Natural Gas	0.07		0.06		0.06		0.07	
Power	0.00		(0.01)		(0.00)		(0.01)	
CO ₂ TS&M	0.06		0.05		0.05		0.06	
Total^B	1.46	1.72	1.21	1.40	1.24	1.45	1.41	1.66
Cost of CO₂ Captured, \$/tonne CO₂^{BC}	18.8	20.9	18.9	21.1	18.7	21.0	19.0	21.1
Cost of CO₂ Avoided, \$/tonne CO₂^{BD}	29.8	31.9	29.9	32.1	29.7	32.0	30.0	32.1

^A Capacity factor assumed to be 90 percent

^B Values shown are for two financial structures

The first (lower value) is based on the loan guarantee finance structure

The second (higher value) is based on the commercial fuels finance structure

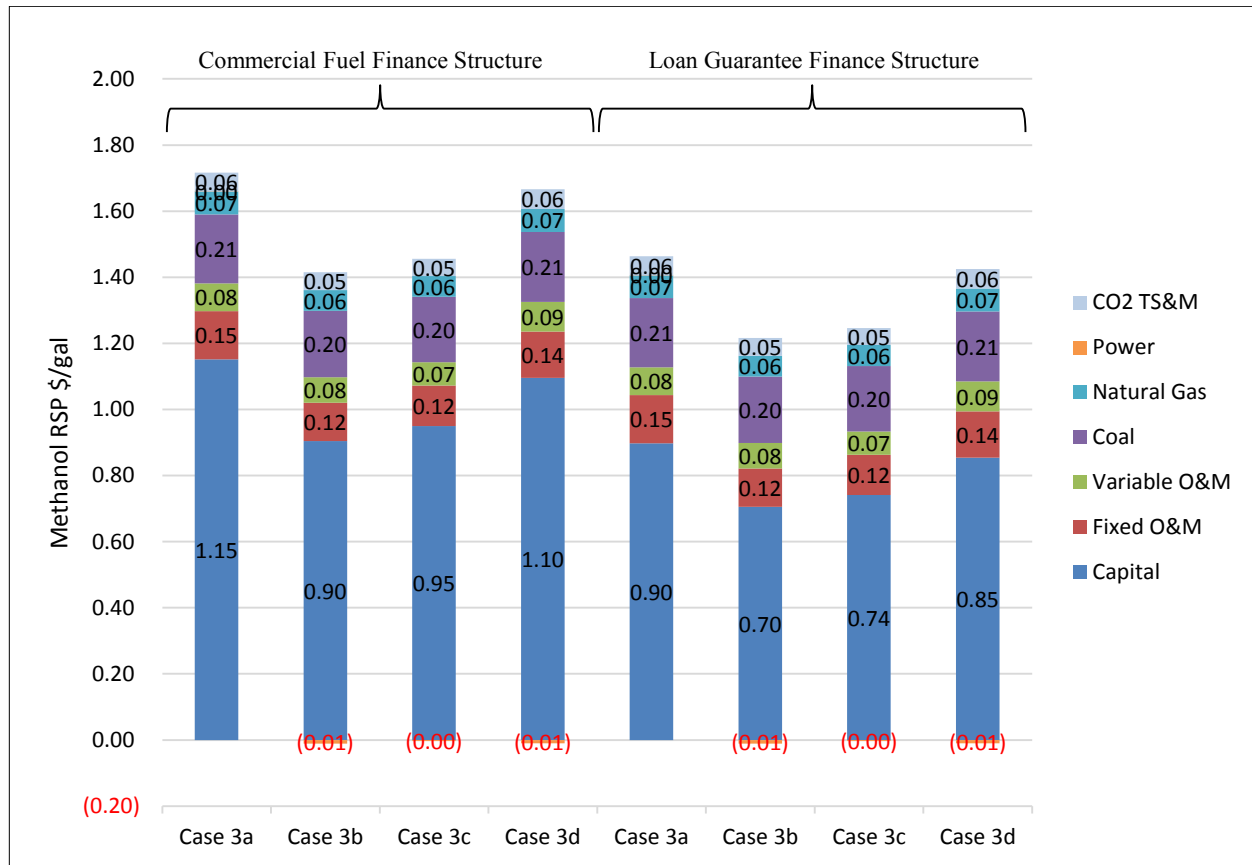
^C Excludes CO₂ TS&M

^D Includes CO₂ TS&M

^E Based on 301.73 gallons/ton

Figure 8-1 illustrates the breakdown of the cost of production. It could be seen from Figure 8-1 that the cases differ most significantly in terms of capital cost on a \$/gal basis.

Figure 8-1
Components of Methanol RSP



This breakdown by component of the methanol RSP for the different cases illustrates the relative contribution of each component. In all cases, the capital cost accounts for 60% or more of the methanol RSP. Both the AR gasification and RTI advanced warm syngas cleanup significantly reduce capital cost and methanol RSP. The AR gasification system reduces the capital cost of the overall gasification block (including the dry feed system) by 40-50% and results in about \$855MM or an 18% decrease in TPC of the entire CTM plant (including benefits to other blocks due to efficiency improvements). The RTI advanced warm syngas cleanup system reduces capital costs of the overall gas cleanup block (including AGR, SRU, and WGS,) by 30-40% and results in reduction in TPC of the entire CTM plant by about \$280MM or about 6%. These capital cost reductions result in a 16% and 4% decrease in methanol RSP for the AR gasification and RTI advanced warm syngas cleanup technologies, respectively, for both financial structures.

The inherent ability of the AR DSP and gasification systems to operate at higher pressures leads to a number of synergistic advantages that result in additional and more general capital cost reductions. The benefits include:

- Elimination of a syngas compressor to boost the syngas pressure for the methanol synthesis process,
- Reduction of the capital cost for the two-stage Rectisol® AGR, and

-
- General reduction in equipment size throughout the syngas generation and cleanup processes associated with the reduced volumetric flow at higher pressures.

An additional benefit of RTI's advanced warm syngas cleanup is the reduction in sulfur emissions. The sulfur emissions for Case 3a (reference case) and 3c (AR gasifier and Rectisol® AGR) are 0.035 and 0.040 lb of SO₂/MMBTU of coal, respectively. For both cases based on RTI's advanced warm syngas cleanup (Cases 3b and 3d), the sulfur emissions are <0.0001 lb of SO₂/MMBTU of coal. This represents over a 99% reduction in sulfur emissions vs the Base Case 3a. The cases with RTI advanced warm syngas cleanup also reduce net CO₂ emissions by between 380 tpd and 420 tpd, which is equivalent to about a 13-15% reduction in CO₂ emissions.

The Case 3a reference plant consumes about 0.8 MW of net power. Because of auxiliary load reductions and/or improved steam generation from process heat recovery, the case studies with AR and/or RTI advanced technologies result in net power generation. For both cases with RTI's advanced warm syngas cleanup, sufficient steam is recovered from the process to generate ≥ 24 MW of net power. This illustrates the benefits of the higher overall process efficiencies enabled by the RTI's WDP process. Case 3c with AR's gasification system generates about 8 MW of net power.

The various CTM plant cases show that at the lower operating pressure used in Cases 3a (reference case) and 3d (Shell gasifier with RTI's advanced warm syngas cleanup), auxiliary loads associated with the chemical solvent used in AACRP are less than those associated with the physical solvent used in Rectisol®. At the higher operating pressure enabled by AR DSP technology, the auxiliary load for the physical solvent-based process (Rectisol®) becomes less than for the chemical solvent based process (AACRP). Although the difference is less than 6% of the total auxiliary load for the CTM plant, this difference illustrates the importance of considering operating pressure in the criteria for CO₂ capture technology selection. In this case, it might have been interesting to see how a coupling of RTI WDP with Rectisol® would have differed from WDP with AACRP.

Because of the increased amount of CO₂ capture for the cases with RTI advanced warm syngas cleanup, the auxiliary loads associated with the CO₂ compression are higher. The overall reductions in auxiliary loads and increased power generated from process heat recovery for the cases with RTI advanced warm syngas cleanup result in over 24 MWe of net power generation, which more than offsets the increased auxiliary load associated with the additional CO₂ compression.

The magnitudes of the cost advantages found in this study reflect the specific constraints associated with this study. The reference case selected for this study was based on Case 2 in the Crude Methanol Study. One of the key design constraints in Case 2 was the decision for the overall CTM plant to be essentially power neutral and for any power generated to also include >90% carbon capture on its emissions. This decision, while allowing for a clear demonstration of >90% carbon capture for the methanol product, forces the inclusion of a dedicated NGCC power plant with carbon capture. This has several very negative impacts. If any excess power is generated, this excess power would probably be sold to the grid at a discount because any

negotiated power purchase agreement would exploit the fact that the power production is an inherent by-product of the methanol process. The breakdown of the methanol RSPs and sensitivity analysis demonstrate that any excess power generation has essentially no financial value, whereas a decision in the base case to allow power needs to be purchased across the fence from the grid would have enabled the net power generated by the AR and RTI advanced technologies to offset power purchases and reduce the methanol RSP. The requirement to include power generation processes also increases the overall capital cost of the CTM plant. As the capital cost represents over 60% of the methanol RSP, the cost of the power generation system and supporting subsystems adversely inflate the capital cost of the CTM plant and methanol RSP and somewhat diminish the relative net impact of utilizing the AR and RTI advanced technologies. Unfortunately sectioning out the capital cost of the power generating equipment and the associated increase in capital cost of the supporting systems is not possible in the current study, because of integration between the process heat recovery and power generation system. Finally, the inclusion of power generation processes also increases the overall auxiliary load by about 8%.

In Case 2, the CTM plant has a total of 9 Shell gasifiers with eight in continuous operation and one spare. Based on the coal throughput, the size of these Shell gasifiers is about 2,000 tpd. Based on the assumption that the reference case should reflect actual commercially operating plants, this gasifier selection is practical. However, this study focuses on evaluating emerging technologies which are expected to be demonstrated commercially in the next 5 years. Because of this approach, the AR gasification system included a 3,600 tpd coal gasifier. Shell's webpage also announces that they have designed and will supply a Shell gasifier capable of 3,200 tpd of coal. If Shell's larger gasifier had been used in the reference case, a total of 5 continuously operating gasifiers would have been required, which is identical to the number of AR gasifiers used in this study. There is no doubt that use of the AR gasification system would still result in a substantial reduction in capital cost (because AR gasifiers are estimated to cost substantially less than equivalent-sized Shell gasifiers), but the total magnitude of any cost reduction would be somewhat smaller.

As mentioned above, synergistic benefits from generation and processing the syngas at higher pressure result in additional capital cost reductions. From the perspective of reliable and safe operation of the gasifier, the dry solids feed system is critical. Even momentary inconsistencies in the coal feed into the gasifier will result in transitions between reducing and oxidizing conditions, which would not allow safe operation of the gasifier. Therefore, demonstration of AR's DSP technology will be an important requirement for commercial deployment of this technology. To help account for this potential barrier to commercialization of this technology, a process contingency of 20% has been included for the AR DSP technology.

As part of this study, Nexant estimated the performance and costs of the Rectisol[®] process operating at 815 psia versus 515 psia. It was determined that the Rectisol[®] process' cost as reported in the NETL Crude Methanol Study is within the lower end of its range of cost estimates. Therefore, the cost savings from utilizing RTI's advanced syngas cleanup system reported in this study are in all likelihood conservative.

The importance of operating pressure on the cost and performance of the CO₂ capture system was identified. Because RTI's advanced syngas cleanup system is integrated with the upstream gasification processes and the downstream methanol synthesis processes, there are many constraints that must be met. Although all of the heat and mass balance constraints were met, optimizing the pressure drop across this system was not a priority. As a consequence, additional optimization of the pressure drop across the RTI advanced syngas cleanup system could further increase economic performance benefits for this system.

Since one of the key benefits of RTI's WDP technology is that it decouples the sulfur and CO₂ removal processes, the selection of the CO₂ capture technology should be made based on economic and performance criteria. For integration of RTI's advanced syngas cleanup system and CO₂ capture in this study, an AACRP system was chosen based on the cost data for an integrated system available from RTI's 50 MWe demonstration plant project (DE-FE0000489). Although this did take advantage of the best available cost information, it did not necessarily maximize the benefits of RTI's advanced syngas cleanup technology. In a future study, the potential benefits of integration of RTI's advanced syngas cleanup technology with Rectisol® for carbon capture would be an interesting case study, particularly for chemical production applications that would likely benefit from higher syngas pressures.

The substantial benefits from integration of AR's gasification and RTI's advanced syngas cleanup into a CTM plant, which include significant capital cost reductions, lower sulfur and CO₂ emissions, and improved energy efficiency, will generate market forces that will drive the implementation of the AR and RTI technologies to realize these benefits. The sensitivity analysis completed on the impact of increasing the estimated capital costs of the AR gasification and the RTI advanced syngas cleanup technologies by as much as 30% still demonstrated significantly lower methanol RSPs. All these factors indicate that when ready for commercial deployment, these technologies will be favored over existing commercial technologies.

°F	Degree Fahrenheit
AACRP	Activated amine CO ₂ recovery process
AGR	Acid Gas Removal
AOI	Area of Interest
AR	Aerojet Rocketdyne
Ar	Argon
ASU	Air Separation Unit
B/L	Battery Limit
BACT	Best Available Control Technology
BEC	Bare Erected Cost
BFD	Block Flow Diagram
BFW	Boiler Feed Water
BOP	Balance of Plant
Btu	British Thermal Unit
CAPEX	Capital Expenditure
CCF	Capital Charge Factor
CF	Capacity Factor
CH ₄	Methane
Circ	Circulating
CMT	Constant Maturity Treasury
CO	Carbon Monoxide
CO ₂	Carbon Dioxide
COE	Cost of Electricity
COS	Carbonyl Sulfide
CTL	Coal-to-liquids
CTM	Coal-to-Methanol
CTG	Combustion Turbine Power Generation
CW	Cooling Water
DBT	Dry Bulb Temperature
DCS	Distributed Control System
DOE	U.S. Department of Energy
DSP	Dry Solids Pump
DSRP	Direct Sulfur Reduction Process
EPC	Engineering, Procurement and Construction
EPRI	Electric Power Research Institute
FCC	Fluid Catalytic Cracking
FO	Fuel Oil

FOA	Funding Opportunity Announcement
ft	feet
gal	Gallon
GE	General Electric
H ₂	Hydrogen
H ₂ O	Water
H ₂ S	Hydrogen Sulfide
Hg	Mercury
HGCU	Hot Gas Clean Up
HHV	Higher Heating Value
HMB	Heat and Material Balance
HP	High Pressure
hr	Hour
HRSG	Heat Recovery Steam Generator
I & C	Instrumentation & Control
IGCC	Integrated Gasification Combined Cycle
IOU	Investor Owned Utility
IP	Intermediate Pressure
kWe	Kilowatt electric
kWh	kilowatt hour
LAER	Lowest available emissions rate
lb	Pound Mass
LHV	Lower Heating Value
LIBOR	London Interbank Offered Rates
LNB	Low NO _x Burner
LOX	Liquid oxygen
LP	Low Pressure
LTGC	Low temperature gas cooling
max	Maximum
ME	Major Equipment
MEC	Major Equipment Cost
MeOH	Methanol
min	Minimum
Misc	Miscellaneous
MM	million
MU	Makeup
MWe	Megawatt electric
MWh	megawatt hour
N ₂	Nitrogen
NETL	National Energy Technology Laboratory

NGCC	Natural Gas Combined Cycle
NO _x	Oxides of Nitrogen
NSR	New Source Review
O&M	Operating and Maintenance
O ₂	Oxygen
OPEX	Operating Expenditure
OSBL	Outside Battery Limit
PC	Pulverized Coal
PCC	Post-Combustion Capture
PFB	Pressurized Feed Bin
PFD	Process Flow Diagram
PM	Particulate Matter
ppmv	Parts per Million by Volume
ppmW, ppmw	Parts per Million by Weight
PRB	Powder River Basin
PSFM	Power Systems Financial Model
psi	Pounds Per Square Inch
psia	Pounds Per Square Inch, absolute
psig	Pounds Per Square Inch, gauge
QGESS	Quality Guidelines for Energy System Studies
RTI	Research Triangle Institute
RSP	Required Selling Price
SC	Supercritical
SCGP	Shell Coal Gasification Process
scfm	Standard Cubic Feet per Minute
SO ₂	Sulfur Dioxide
SOPO	Statement of Project Objectives
SRU	Sulfur Recovery Unit
STG	Steam Turbine Power Generation
T&S	Transportation and Storage
TDC	Total Direct Cost
TEA	Techno-Economic Analysis
TFC	Total Field Cost
TG	Turbine Generator
TGTU	Tail Gas Treatment Unit
TIC	Total Installed Cost
TOC	Total Overnight Cost
TPC	Total Plant Cost
tpd	tons per day
US, USA	United States of America

vol%	Percentage by Volume
WBT	Wet Bulb Temperature
WDP	Warm Gas Desulfurization Process
WGS	Water Gas Shift
WHR	Waste Heat Recovery
WT	Waste Treatment

12. Appendix C

Benefits of Integration of Aerojet Rocketdyne and RTI Advanced Gasification Technologies for Hydrogen-Rich Syngas Production

Task 5: Techno-Economic Analysis for Coal-to-Liquids via Fischer-Tropsch Synthesis

Project Period:
October 1, 2013 to September 30, 2016

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This study evaluated the techno-economic benefits of integrating the advanced compact gasification technology being developed by Aerojet Rocketdyne (AR, currently being offered by Gas Technology Institute as R-GAS™) and the advanced warm gas desulfurization process technology developed by Research Triangle Institute (RTI), for coal-to-liquids (CTL) via a cobalt-based Fischer-Tropsch (FT) synthesis process. RTI's advanced syngas cleanup process consists of the following components: RTI's warm gas desulfurization process (WDP), RTI's direct sulfur recovery process (DSRP), and advanced fixed-bed water gas shift process (AFWGS). An activated amine CO₂ recovery process (AACRP) is also integrated with RTI's advanced syngas cleanup process to effectively remove CO₂. AR's compact gasifier is an entrained flow, dry-feed gasification system that utilizes a dry solids pump (DSP) as the coal feed system. These integrated advanced technologies were found to offer significant benefits relative to a U.S. Department of Energy (DOE) state-of-the-art reference case (Shell gasifier with Illinois #6 coal feed, plus Rectisol® acid gas removal technology). Both the AR gasification and RTI advanced warm syngas cleanup technologies significantly reduce capital cost and FT diesel required selling price (RSP). The AR gasification system reduces the capital cost of the overall gasification block (including the dry feed system) by 35% and results in reduction in total plant cost (TPC) of the entire CFTL plant by about \$610 MM or a 11.5% decrease. The RTI advanced warm syngas cleanup system reduces capital costs of the overall gas cleanup block (including acid gas removal, sulfur recovery, syngas cooling, and water gas shift) by 24% and results in reduction in TPC of the entire CFTL plant by \$211 MM or about 4%. These capital cost reductions result in a 10% and 3% decrease in the FT diesel RSP for the AR gasification and RTI advanced warm syngas cleanup technologies, respectively. The combined advanced technologies yield a 15% reduction in overall CFTL capital cost and a 13% lower RSP of FT diesel liquids compared to the reference case. It should be noted that the DOE reference case, and thus our comparison cases, were based on use of high-chlorine Illinois #6 coal, which produces substantial levels of HCl in the raw syngas. Since high levels of HCl in the syngas are detrimental to all the downstream chemical processes, including WDP, this required the addition of a process for removal of the HCl. In the reference case, HCl was removed in the syngas scrubber and further treatment of the dissolved chlorine was completed during waste water treatment. To optimize thermal efficiency, a warm-temperature non-regenerable HCl guard bed was used to remove HCl upstream of WDP. The factored cost estimate used to cost the balance of the plant processes, which includes water treatment, prohibited removing an additional cost associated with treatment of the dissolved chlorine. Thus, the cases with WDP include the cost for treatment of the HCl twice. This means that the resulting capital cost and RSP for these specific WDP cases are overestimated and for most other coals, which have low chlorine, the benefits for use of WDP would be significantly better.

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Executive Summary

With DOE funding from Cooperative Agreement DE-FE0012066, Research Triangle Institute (RTI), Aerojet Rocketdyne (AR), and Nexant formed a team to evaluate the cost and performance benefits associated with integration of the advanced gasification technologies being developed by AR and the warm gas cleanup technologies being developed by RTI for integrated gasification combined-cycle (IGCC) and coal-to-liquids (CTL) applications with $\geq 90\%$ carbon capture. Work on IGCC and coal-to-methanol (CTM) cases was performed by Nexant and RTI while the work on the coal-to-Fischer-Tropsch-liquids (CFTL) cases was performed by RTI. This report documents the study completed for CFTL. Separate reports have been prepared for the IGCC and CTM studies.

RTI's warm gas desulfurization process and AR's advanced gasification system are next generation technologies that will enable commercial deployment of gasification through improved efficiency, greater contaminant removal, and lower costs within the next five years. AR's compact gasifier (currently being offered by Gas Technology Institute as R-GASTM) is an entrained flow, dry-feed gasification system that utilizes a dry solids pump (DSP) as the coal feed system. Due to its novel design, the gasifier is drastically smaller (90% size reduction) than a conventional gasifier. Used in conjunction with the dry solids pump, which replaces a large and costly lockhopper-based pressurized coal feed system, the overall compact gasification system is projected to cost as little as 50% of that of a conventional gasification system.

RTI's advanced syngas cleanup process consists of the following components: RTI's warm gas desulfurization process (WDP), RTI's direct sulfur recovery process (DSRP), and advanced fixed-bed water gas shift process (AFWGS). An activated amine CO₂ recovery process (AACRP) is also integrated with RTI's advanced syngas cleanup process to effectively remove CO₂. This combination effectively decouples the separation of the sulfur and CO₂ enabling more efficient removal of these contaminants at lower cost. Because this combination is also able to produce syngas with contaminant concentrations similar to a Rectisol[®] process, RTI's advanced syngas cleanup process offers the potential for significant reduction in contaminant emissions at lower cost than commercially available technologies. With RTI's 50 MW_e demonstration plant project (DE-FE0000489) having demonstrated the technology at Tampa Electric Company's Polk Power Station, RTI's advanced syngas cleanup technology is now ready for commercial deployment.

The objective of this study was to assess the benefits of integrating the technologies offered by RTI and AR for reducing the capital and production costs for syngas generation and cleanup in a CFTL plant with $\geq 90\%$ carbon capture.

The reference case for this study uses a Shell gasifier to convert Illinois #6 coal into FT liquids using a water gas shift reactor to alter the H₂/CO ratio of the raw syngas, a Rectisol[®] syngas cleanup process, a cobalt-based FT process with product upgrading to produce the final fuel products, and a combined cycle system to recover the auxiliary power from the hydrocarbon byproduct. This case was developed by combining the information from different U.S. Department of Energy (DOE) studies on coal-to-liquids using a Shell gasifier with an iron-based catalyst, and a gas-to-liquids study using a Co-based catalyst.

To fully evaluate the benefits of integrating these technologies, a total of three case studies were completed. The specific technologies included in each of the three CFTL plant configurations are identified in the CFTL case study matrix shown in Table 0-1.

Table 0-1
Case Study Matrix for Coal-to-FT-Liquids Production with CO₂ Capture

	Case 5A ¹	Case 5B	Case 5C
Gasification Technology			
Shell Gasification	✓	✓	
AR Gasification Technology			✓
Gas Cleanup			
Rectisol [®] for CO ₂ and Sulfur Removal ²	✓		
RTI WDP with Activated Amines CO ₂ Recovery Process (AACRP)		✓	✓
Water Gas Shift			
Sour Shift	✓		
Advanced Fixed Bed Water-Gas Shift		✓	✓
FT Synthesis			
Conventional technology ³	✓	✓	✓
CO ₂ Drying and Compression (to 2,215 psia)	✓	✓	✓

¹ Case 5A is the reference case. Gasifier selection depends on selected DOE NETL baseline studies.

² Rectisol[®] remove sulfur and CO₂. Selection and any additional trace contaminant cleanup technologies needed will be as defined by DOE NETL baseline studies.

³ Baseline FT technology will be established based on DOE NETL guidelines/recommendations

Table 0-2 shows a summary comparison of the capital costs, operating costs, FT liquids production, plant power import/export, auxiliary power consumption, required selling price (RSP) of FT diesel liquids, and cost of CO₂ captured/avoided for the three cases.

Table 0-2
CTL Results Summary

Case	Case 5A		Case 5B		Case 5C	
Gasifier	Shell		Shell		Aerojet	
Sulfur and CO ₂ Removal	Dual Stage Rectisol [®]		RTI WDP/AACRP		RTI WDP/AACRP	
CAPEX, \$MM						
Total Installed Cost (TIC)	\$3,882		\$3,734		\$3,220	
Total Plant Cost (TPC)	\$5,405		\$5,187		\$4,577	
Total Overnight Cost (TOC)	\$6,771		\$6,530		\$5,797	
OPEX, \$MM/yr (100% Capacity Factor Basis)						
Fixed Operating Cost (OC _{fix})	\$205.8		\$198.9		\$179.8	
Variable Operating Cost, less Fuel (OC _{var})	\$176.1		\$190.0		\$179.3	
Coal Fuel (OC _{fuel})	\$526.0		\$526.0		\$526.0	
Import/(Export) Power (OC _{Power})	(\$21.7)		(\$36.9)		(\$30.7)	
Total OPEX	\$886.1		\$878.0		\$854.3	
Plant Output						
Crude FT Diesel Product, bbl/day	34,992		34,994		36,115	
Crude FT Naphtha Product, bbl/day	15,008		15,006		15,486	
Net Power Output, MWe	41.3		70.2		58.4	
Power Generation						
Gas Turbine Power	122.5		101.5		104.9	
Steam Turbine Power	403.2		448.6		414.2	
Syngas Turboexpansion Power	7.3		10.0		11.4	
Total Gross Power	533.0		560.0		530.4	
Net Power Generation/(Consumption), MWe	41.3		70.2		58.4	
Common Factors Used for RSP Calculation						
Capital Charge Factor (CCF)	0.1700	0.218	0.1700	0.218	0.1700	0.218
Capacity Factor (CF)	0.9		0.9		0.9	
Required Selling Price^{A,B,C}						
RSP FT Diesel, \$/bbl _{FTD}	134.3	156.0	131.0	152.0	117.3	135.6
RSP FT Naphtha, \$/bbl _{FTN}	94.2	109.2	89.9	104.6	80.4	93.1
Equivalent Crude Oil Price, \$/bbl _{ECO}	116.5	135.4	113.7	131.9	101.8	117.7
Equivalent Petroleum Diesel Price, \$/bbl _{EPD}	145.7	169.3	142.1	164.9	127.3	147.1
F-T Diesel Required Selling Price Breakdown						
CAPEX	77.02	98.76	74.27	95.25	63.89	81.93
Fixed O&M	13.77	13.77	13.31	13.31	11.65	11.65
Variable O&M	10.60	10.60	11.44	11.44	10.46	10.46
Fuel	31.67	31.67	31.67	31.67	30.69	30.69
Power Generation	-1.31	-1.31	-2.22	-2.22	-1.79	-1.79
CO ₂ T&S	2.50	2.50	2.52	2.52	2.39	2.39
Total	134.25	156.00	131.00	151.97	117.29	135.33
CO₂ Capture						
CO ₂ Capture Product, tpd	25,116		25,280		24,729	
Carbon Captured (as % Carbon in Fuel)	92.5%		92.8%		92.6%	
Costs of CO₂ Captured^{A,B,C} (\$/ton)	13.92	15.00	13.44	14.49	13.47	15.04

^A Capacity factor assumed to be 90 percent

^B Values shown are for two financial structures

The first (lower value) is based on the loan guarantee finance structure

The second (higher value) is based on the commercial fuels finance structure

^C Includes CO₂ T&S

Integration of the AR gasification technology and RTI's advanced syngas cleanup technologies into a CFTL plant offers significant cost reductions, and contributes to reducing the RSP of the

FT diesel and FT naphtha. The technoeconomic analysis and the results show the substantial benefits from integration of AR's gasification and RTI's advanced syngas cleanup into a CFTL plant, which include capital cost reductions and improved energy efficiency.

Both the AR gasification and RTI advanced warm syngas cleanup technologies significantly reduce capital cost and FT diesel RSP. The AR gasification system reduces the capital cost of the overall gasification block (including the dry feed system) by 35% and results in reduction in total plant cost (TPC) of the entire CFTL plant by about \$610 MM or a 11.5% decrease. The RTI advanced warm syngas cleanup system reduces capital costs of the overall gas cleanup block (including AGR, SRU, and WGS,) by 24% and results in reduction in TPC of the entire CFTL plant by \$211 MM or about 4%. These capital cost reductions result in a 10% and 3% decrease in the FT diesel RSP for the AR gasification and RTI advanced warm syngas cleanup technologies respectively, for both financial structures. The combined advanced technologies yield a 15% reduction in overall CFTL capital cost and a 13% lower RSP of FT diesel liquids compared to the reference case.

It should be noted that the DOE reference case and thus our comparison cases were based on use of high-chlorine Illinois #6 coal, which produces substantial levels of HCl in the raw syngas. Since high levels of HCl in the syngas are detrimental to all the downstream chemical processes, including WDP, this required the addition of a process for removal of the HCl. In the reference case, HCl was removed in the syngas scrubber and further treatment of the dissolved chlorine was completed during waste water treatment. To optimize thermal efficiency, a warm-temperature non-regenerable HCl guard bed was used to remove HCl upstream of WDP. The factored cost estimate used to cost the balance of the plant processes, which includes water treatment, prohibited removing any additional cost associated with treatment of the dissolved chlorine. Thus, the cases with WDP include the cost for treatment of the HCl twice. This means that the resulting capital cost and RSP for these WDP cases are overestimated and for most other coals, which have low chlorine, the relative economic benefits for Cases 5B and 5C reported in Table 0-2 above would be increased.

1.1 BACKGROUND

With DOE funding from Cooperative Agreement DE-FE0012066, Research Triangle Institute (RTI), Aerojet Rocketdyne (AR), and Nexant formed a team to evaluate the cost and performance benefits associated with integration of the advanced gasification technologies being developed by AR and the warm gas desulfurization process being developed by RTI for integrated gasification combined cycle (IGCC) and coal-to-liquids (CTL) applications with $\geq 90\%$ carbon capture. Work on coal-to-methanol (CTM) case studies was performed by Nexant and RTI while the work on the coal-to-Fischer-Tropsch-liquids (CFTL) was performed by RTI.

RTI's warm gas desulfurization process and AR's advanced gasification system are next generation technologies that will enable commercial deployment of gasification through improved efficiency, greater contaminant removal, and lower costs within the next five years. AR's compact gasifier (currently being offered by Gas Technology Institute as R-GASTM) is an entrained flow, dry-feed gasification system that utilizes a dry solids pump (DSP) as the coal feed system. Due to its novel design, the gasifier is drastically smaller (90% size reduction) than a conventional gasifier. Used in conjunction with the dry solids pump, which replaces a large and costly lockhopper-based pressurized coal feed system, the overall compact gasification system is projected to cost as little as 50% of that of a conventional gasification system.

RTI's advanced syngas cleanup process consists of the following components: RTI's warm gas desulfurization process (WDP), RTI's direct sulfur recovery process (DSRP), and advanced fixed-bed water gas shift process (AFWGS). An activated amine CO₂ recovery process (AACRP) is also integrated with RTI's advanced syngas cleanup process to effectively remove CO₂. This combination effectively decouples the separation of the sulfur and CO₂ enabling more efficient removal of these contaminants at lower cost. Because this combination is also able to produce syngas with contaminant concentrations similar to a Rectisol[®] process, RTI's advanced syngas cleanup process offers the potential for significant reduction in contaminant emissions at lower cost than commercially available technologies. With RTI's 50 MWe demonstration plant project (DE-FE0000489) having demonstrated the technology at Tampa Electric Company's Polk Power Station, RTI's advanced syngas cleanup technology is now ready for commercial deployment.

1.2 STUDY OBJECTIVES

The overall objective of this project is to assess the potential of integrating the technologies offered by RTI and AR to reduce the capital and production costs for hydrogen-rich syngas with near-zero emissions from coal gasification for both power and CTL productions with carbon capture. To fully evaluate the integration of these advanced technologies, separate studies were completed for hydrogen-rich syngas production for power and coal-to-liquids applications.

This report documents the study completed for coal-to-liquids via Fischer-Tropsch (FT) synthesis. Separate reports have been prepared for power production via IGCC and coal-to-methanol studies.

The specific case studies completed provide a comparison of each advanced technology individually and as an integrated plant with a reference plant using commercially available

technologies. All comparison studies conducted for this report capture at least 90% of the carbon in the feedstock as either CO₂ for storage or in the liquids product.

2.1 DESIGN REFERENCES

Several DOE baseline studies for reference plants exist for the coal-to-FT liquids (CFTL) process. There are two studies that use coal gasification for syngas generation and an FT reactor system using an iron-based catalyst. The “Baseline Technical and Economic Assessment of a Commercial Scale Fischer-Tropsch Liquids Facility, April 9, 2007, DOE/NETL-2007/1260” (CTL Study 1) [1] uses an Illinois #6 bituminous coal and the ConocoPhillips E-Gas gasification technology and Selexol for acid gas removal upstream of the FT process. The “Cost and Performance Baseline for Fossil Energy Plants. Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis, October 15, 2014, DOE/NETL-2011/147” (CTL Study 2) [2] also uses the Illinois #6 bituminous coal, but uses the Shell gasification technology and Rectisol® for acid gas removal.

Since the iron-based catalyst is also active for the water gas shift reaction, a separate water gas shift process is not required upstream of the FT process. Both water gas shift and FT-synthesis occur in the FT-synthesis plant. Majority of the CO₂ capture thus occurs downstream of the FT process. As these CO₂ capture processes are operated at lower pressure due to their location in the overall process, CO₂ capture efficiency is reduced.

Another study, the “Analysis of Natural Gas-to-Liquid Transportation Fuels via Fischer-Tropsch”, September 13, 2013, DOE/NETL-2013/1597 (GTL Study) [3], uses the Co-based catalyst for the FT process. In the GTL study, natural gas is used to generate syngas with a H₂/CO ratio of 2.0-2.2, and sent to the FT-reactor that utilizes a Co-based catalyst for FT synthesis. The FT product upgrading in this study is more extensive than the CTL study and provides roughly 30:70 split of gasoline and diesel fuel.

In addition to these recent DOE studies, there are some reports from Bechtel Corporation on DOE-sponsored work on Fischer-Tropsch liquids production using different feedstocks [4-7]. These reports provide extensive details on the FT process for both the iron-based and the cobalt-based catalyst, with limited details on the syngas generation and acid gas cleanup.

An ideal reference case for use as a comparison case for the CFTL is not readily available. RTI created a reference case from existing DOE case studies by extracting process components from the existing case studies and combining them. The reference plant design used for the coal-to-FT liquids (CFTL) study was selected as a combination of two DOE case studies, “Cost and Performance Baseline for Fossil Energy Plants. Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis, October 15, 2014, DOE/NETL-2011/147” (CTL Study 2) [2] and the “Analysis of Natural Gas-to-Liquid Transportation Fuels via Fischer-Tropsch”, September 13, 2013, DOE/NETL-2013/1597 (GTL Study) [3].

By combining the information from different DOE studies, a reference case was developed that uses a Shell gasifier to convert Illinois #6 coal into FT liquids using a water gas shift reactor to alter the H₂/CO ratio of the raw syngas, dual-stage Rectisol® as a single syngas cleanup process, a cobalt-based FT process with product upgrading to produce the final fuel products, and a combined cycle system to recover the auxiliary power from the hydrocarbon byproduct. This reference case is expected to provide significant benefits of having a single CO₂ capture upstream of the FT reactor. The savings in a single CO₂ capture process at higher pressures is

expected to out-weigh the differential costs for a sour water-gas shift process and for the cobalt-based catalyst versus the iron-based catalyst.

The FT-liquids production from syngas was modeled after the DOE sponsored study on natural gas to FT liquids model by Bechtel [4]. As the Bechtel models form the basis for all subsequent DOE studies, the Bechtel model using the Co-based catalyst [4] was used to model the FT process.

This CFTL case was used to establish a viable commercial reference and to compare the impacts of integration of the AR gasification and RTI advanced syngas cleanup and AACRP systems. In addition, the assumptions and calculation methodologies described in the following DOE/NETL's Series of Quality Guidelines for Energy Systems Studies (QGESS) were used in this work:

- “Specifications for Selected Feedstocks, January 2012, DOE/NETL. 341/011812” [8],
- “Process Modeling Design Parameters, January 2012, DOE/NETL. 341/081911” [9],
- “CO₂ Impurity Design Parameters, January 2012, DOE/NETL. 341/011212” [10],
- “Detailed Coal Specifications, January 2012, DOE/NETL-401/01211” [11],
- “Cost Estimation Methodology for NETL Assessments of Power Plant Performance, April 2011, DOE/NETL. 2011/1455” [12],
- “Capital Cost Scaling Methodology, January 2013, DOE/NETL. 341/013113” [13], and
- “Fuel Prices for Selected Feedstocks in NETL Studies, November 2012, DOE/NETL 341/11212” [14]

“Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases, August 2012, DOE/NETL-341/082312” (NETL Report 341/082312), was used as the reference to develop the updated capital and operating cost estimates, in June 2011 dollars [15].

The report uses the DOE baseline cases as the basis for comparison for the technoeconomic analyses. Since all process descriptions and results other than the technologies under evaluation are taken from DOE baseline studies, a lot of the text from the DOE reports is used verbatim and the format of the result tables mirror those in the DOE reports.

2.2 CASE CONFIGURATIONS

To identify and determine any synergistic advantages of integrating both AR's advanced gasification and RTI's advanced syngas cleanup process, a total of three case studies were completed. The most promising case study was anticipated to be the CFTL plant with CO₂ capture integrating both AR's advanced gasification and RTI's advanced syngas cleanup process (Case 5C). Case 5B provides the evaluation from the use of the RTI advanced syngas cleanup process on a standalone basis.

The three CFTL production plant configurations are identified in the case study matrix shown in Table 2-1.

Table 2-1
Case Study Matrix for Coal-to-FT-Liquids Production with CO₂ Capture

	Case 5A ¹	Case 5B	Case 5C
Gasification Technology			
Shell Gasification	✓	✓	
AR Gasification Technology			✓
Gas Cleanup			
Rectisol [®] for CO ₂ and Sulfur Removal ²	✓		
RTI WDP with Activated Amines CO ₂ Recovery Process (AACRP)		✓	✓
Water Gas Shift			
Sour Shift	✓		
Advanced Fixed Bed Water-Gas Shift		✓	✓
FT Synthesis			
Conventional technology ³	✓	✓	✓
CO ₂ Drying and Compression (to 2,215 psia)	✓	✓	✓

¹ Case 5A is the reference case. Gasifier selection depends on selected DOE NETL baseline studies.

² Rectisol[®] remove sulfur and CO₂. Selection and any additional trace contaminant cleanup technologies needed will be as defined by DOE NETL baseline studies.

³ Baseline FT technology will be established based on DOE NETL guidelines/recommendations

Shell gasification was used for Cases 5A and 5B. The Rectisol[®] process was selected for CO₂ and sulfur removal for the base case, due to the stringent sulfur removal requirements for the FT liquids production process. All three case studies used conventional cobalt-based FT production technology for FT synthesis.

2.2.1 Case 5A: Reference CFTL Production Plant with Rectisol[®]-Based Acid Gas Removal (AGR)

Case 5A is the reference plant design used for the CFTL study and was selected as a combination of two DOE case studies, “Cost and Performance Baseline for Fossil Energy Plants. Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis, Revised October 15, 2014, DOE/NETL-2011/1477” (CTL Study) and the “Analysis of Natural Gas-to-Liquid Transportation Fuels via Fischer-Tropsch”, Revised September 13, 2013, DOE/NETL-2013/1597 (GTL Study).

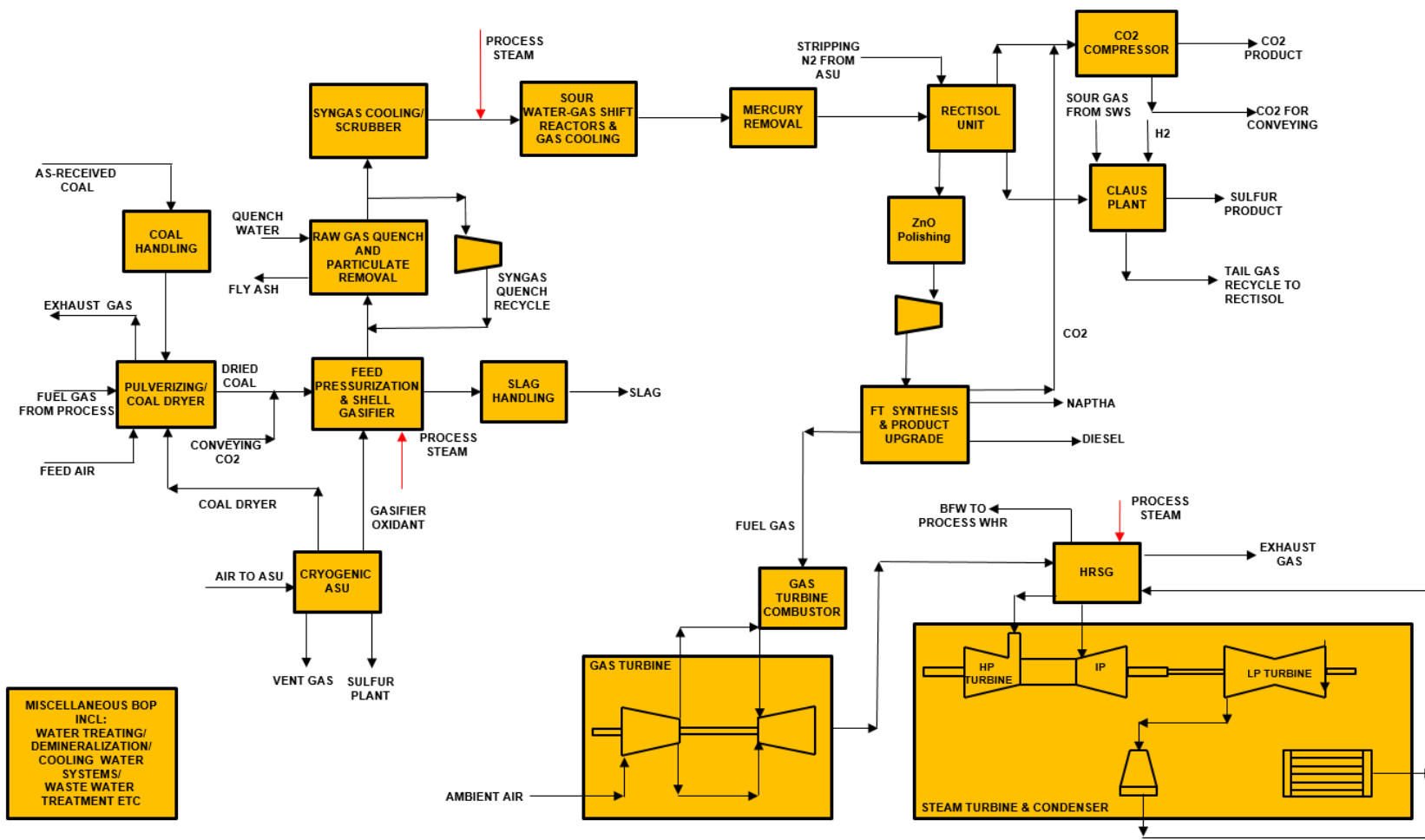
The reference case uses the Illinois #6 bituminous coal and the Shell gasification system is used to generate enough syngas to produce approximately 50,000 barrels per day of FT-liquids with a 30:70 naphtha:diesel split. The size of the plant is consistent with the earlier DOE CTL and GTL studies, which formed the basis for this reference case.

The two-stage Rectisol[®] process was designed for sulfur and CO₂ removal with a target of 90% for CO₂ removal for the overall plant.

Both the gasification and FT-liquids synthesis processes generate a large amount of heat that is recovered, in the form of steam for process requirements and power generation. In combination with the associated steam turbines, the total power generated from the CTL plant's power cycle in the reference case is slightly higher than the total estimated auxiliary loads and the excess power is exported to the grid.

A block flow diagram (BFD) of the Case 5A reference CTL plant is shown in Figure 2-1. It is assumed to operate as a base-loaded unit with an annual on-stream factor of 90% or 7,884 hr/year at full capacity.

Figure 2-1
Case 5A: Reference Shell Gasifier with Rectisol® AGR Coal-to-FT Liquids Plant - Simplified BFD



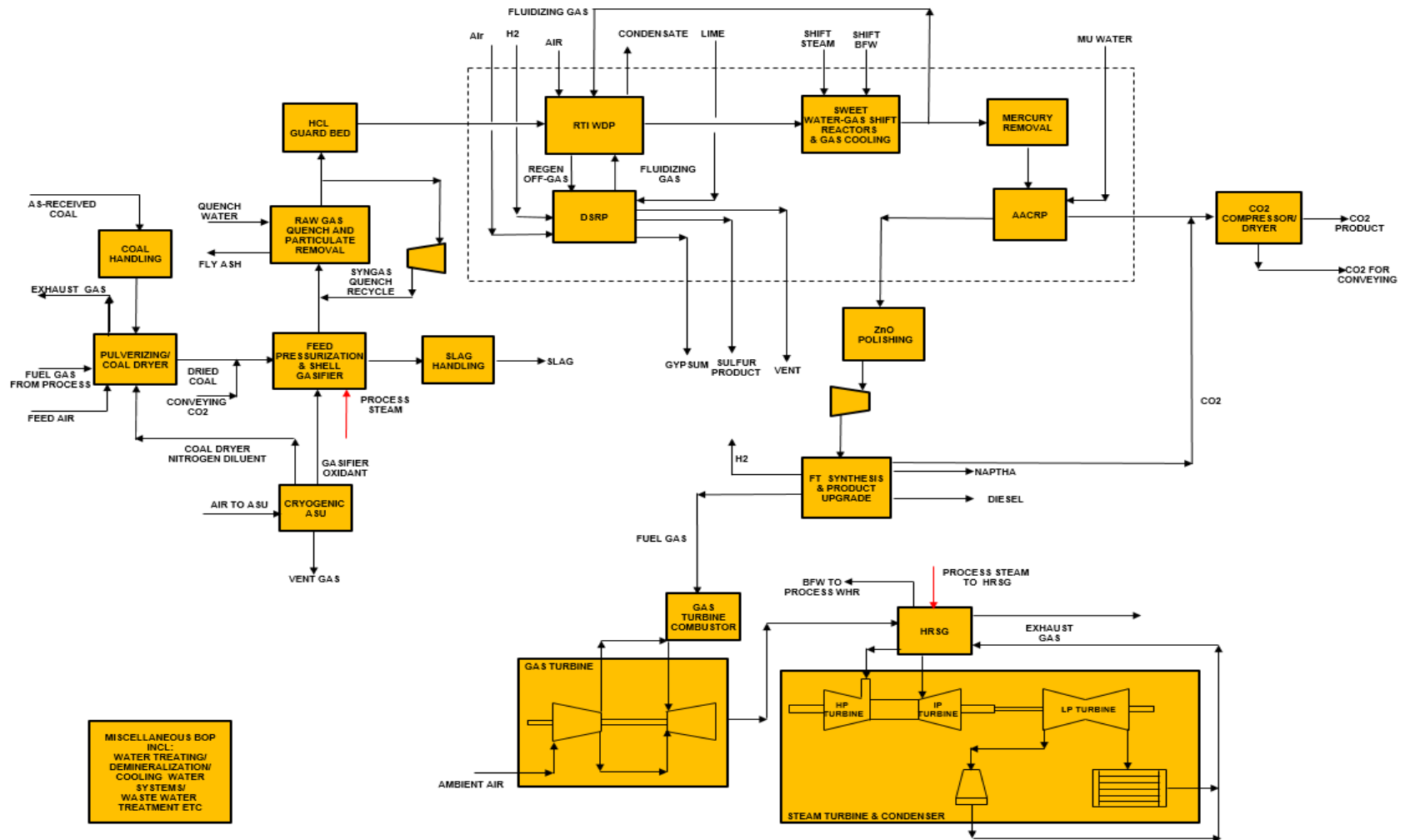
2.2.1 Case 5B: Shell Gasifier with RTI WDP Coal-to-FT-Liquids Plant

Case 5B is a conceptual design that uses the Shell gasification process paired with RTI advanced warm-syngas cleanup and AACRP processes to remove sulfur and CO₂ from the syngas. The coal feed rate is expected to be the same as Case 5A since both utilize Shell gasifiers, which, with the same cold gas efficiencies, generate similar syngas quantities for production of 50,000 barrels per day of FT-liquids (30:70 naptha:diesel) in the FT synthesis units. Three GE MS6001B gas turbines were required to meet the Case 5B plant's auxiliary demands such that no power import is required, with any excess power exported to the grid as byproduct to generate additional revenue.

The blocks within the broken-line rectangle represent RTI's advanced syngas cleanup process and consist of the unit operations designed by RTI. These include the equipment for WDP, DSRP, AFWGS, and the AACRP unit. H₂S and COS are removed from the syngas in the WDP unit while primarily CO₂ is removed via the AACRP unit.

The unchanged blocks consist of the balance of plant processes as in Case 5A but with different throughputs. Their costs are scaled based on system capacities using factors from the QGESS *Capital Cost Scaling Methodology* document.

Figure 2-2
Case 5B: Shell Gasifier with RTI WDP Coal-to-FT-liquids Plant - Simplified BFD



2.2.2 Case 5C: AR Compact Gasifier with RTI WDP Coal-to-FT-Liquids Plant

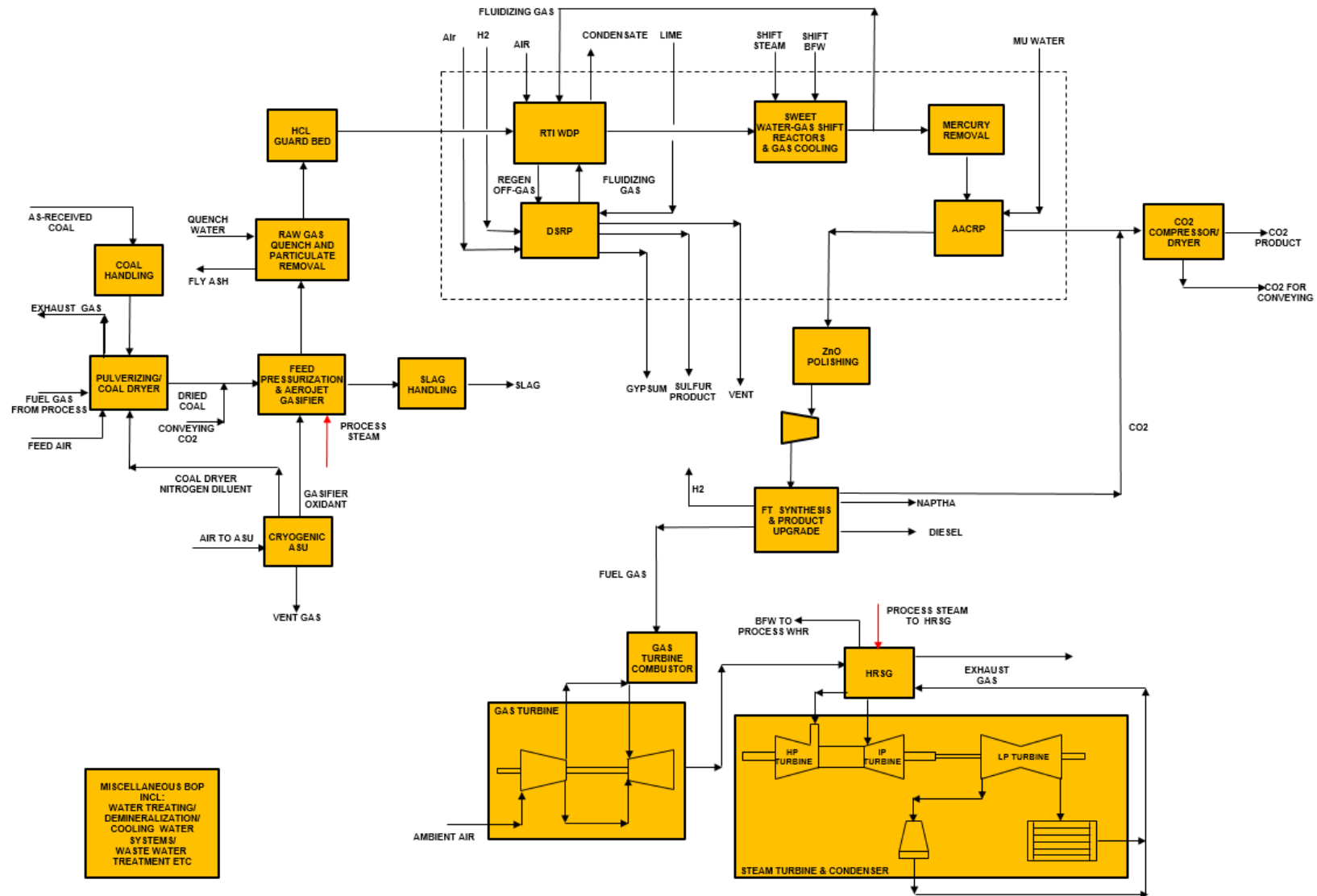
Case 5C is a conceptual design that fully integrates the AR gasification process with the RTI WDP. For power production, three GE MS6001B turbines were required to meet the plant's demands such that no power import is required. Some excess power is produced and exported to the grid for extra revenue. The AR gasifier has a higher cold gas efficiency, and at constant fuel rate, results in excess liquids fuel production. Total FT liquids production in Case 5C is 51,601 bpd, a 3% increase in production.

A combination of RTI's WDP unit and AACRP unit replaces the Rectisol[®] unit in Case 5A to remove the sulfur and CO₂ from the syngas. The Case 5C plant will remove sulfur via RTI's WDP which removes H₂S and COS from the syngas after it leaves the particulate filters, without requiring additional cooling. The treated syngas undergoes sweet shift in high temperature shift process before it is cooled and sent to AACRP unit for CO₂ capture. The AACRP will capture CO₂ equivalent to 90% of the raw syngas' carbon content to meet the CO₂ concentration requirements for FT Liquids synthesis.

The Case 5C plant BFD is shown in Figure 2-3. Here, the AR's coal feed and compact gasifier system, replaces the lockhopper feed system and Shell gasifier in the Case 5A reference plant. The blocks within the broken-line rectangle represent RTI's advanced syngas cleanup and AACRP processes and comprise the WDP for sulfur removal, DSRP to produce elemental sulfur, AFWGS and AACRP for CO₂ removal.

The remaining blocks in Figure 2-3 are the balance of plant processes that are unrelated to the AR gasification and RTI advanced syngas cleanup systems. These blocks have the same unit operations as those in Case 5A, but with different throughputs due to the different flow quantities. Costs for these systems were scaled based on system capacities from Case 5A using factors given in the QGESS Capital Cost Scaling Methodology document wherever possible.

Figure 2-3
Case 5C: AR Compact Gasifier with RTI WDP Coals-to-FT-Liquids Plant - Simplified BFD



2.3 DESIGN CRITERIA

2.3.1 Heat and Material Balance

RTI carried out simulation of all three CTL cases on ASPEN Plus to obtain the process heat and material balances (HMB). This helps to facilitate a more detailed estimation of the overall plant utility balance, including process waste heat recovery, generation from the power cycle, as well as cooling water load breakdown, all of which help to determine the overall CTL plant performance with more certainty.

RTI developed a conceptual design of the Case 5A reference plant with two-stage Rectisol®-based AGR for H₂S and CO₂ removal, and a slurry bubble column reactor with a Co-based catalyst for FT synthesis. Individual models for the different processes were simulated and validated against published DOE baseline studies. The syngas generation using the Shell gasifier with Illinois #6 coal was validated against data from the Case B1B from the 2015 DOE Baseline study [16]. For FT synthesis, earlier DOE sponsored work on GTL studies performed by Bechtel were used [4]. The DOE-sponsored Bechtel studies provide details on the FT product distribution from the FT reactor and product upgrading processes. The Bechtel GTL study uses a Co-based catalyst and the product distribution from that study was used in this work. The RTI model for HRSG and ST cycle was validated against published data from Case B1B from DOE baseline study on bituminous coal for IGCC power [16]. Data from Case B1B was used to determine the steam turbine efficiencies to match the total power and temperature of the steam exiting the turbines. This model was used to determine the ST power in this study. All these validation studies provide confidence that the results from this work are within a reasonable range of accuracy.

2.3.2 AR Compact Gasifier and Feed System

2.3.2.1 DSP Feed System

AR had provided costs for its proprietary dry solids pump (DSP) for the IGCC and CTM studies. The AR DSP feed system has been previously evaluated by Nexant and compared against the Shell lockhopper feed system in a separate study (DE-FE0012062). The results from this study forms the basis for the cost of the DSP system.

2.3.2.2 AR Compact Gasification System

The AR gasifier was modeled in Aspen Plus using the restricted equilibrium approach to determine heat and mass balance and the syngas composition. The model was validated against experimental data from the AR gasifier using Illinois #6 and PRB coal at 1200 psia [17]. The AR gasifier model was integrated into the RTI's CFTL model to determine the syngas flow-rate and composition.

The AR gasifiers used in this study have the same size as used in the earlier IGCC and CTM studies. AR provided the equipment costs for their gasifier for the IGCC and CTM studies. These costs were used to determine the total plant cost (TPC) for the CTL plant for Case 5C.

2.3.3 RTI WDP System

RTI carried out process simulations to perform detailed modeling of its proprietary warm-gas desulfurization process (WDP), the AFWGS and the AACRP to determine the major stream

flows and utilities consumption. The model utilized information from the recently concluded 50 MWe demonstration plant operations. The equipment costs for proprietary equipment were estimated by scale-up of actual cost data from the 50 MWe demonstration plant construction. These were used as inputs to establish the overall CTL plant performance and TPC and variable operating costs for Cases 5A and 5B.

2.3.4 Other Systems

The costs for the balance of plant processes not directly related to the AR gasification, RTI advanced syngas cleanup and AACRP systems were derived via capacity-factor from the corresponding system costs listed in the DOE CTL and DOE GTL baseline studies.

2.4 SITE-RELATED CONDITIONS

The CTL plants in this study are assumed to be located in a generic, Midwestern USA site, with site-related conditions as shown below:

▪ Location	Greenfield, Midwestern USA
▪ Elevation, ft	above sea level 0
▪ Topography	Level
▪ Size, acres	300
▪ Transportation	Rail, Road, Pipeline
▪ Ash/slag disposal	Off Site
▪ Water	Municipal (50%)/Groundwater (50%)
▪ Access	Landlocked, having access by train and highway
▪ CO ₂ disposition	Compressed to 2,215 psia at battery limit

2.5 METEOROLOGICAL DATA

Maximum design ambient conditions for material balances, thermal efficiencies, system design and equipment sizing are:

▪ Barometric pressure, psia	14.696
▪ Dry bulb temperature (DBT)	59 °F
▪ Wet bulb temperature (WBT)	51.5 °F
▪ Ambient relative humidity, %	60

2.6 OTHER SITE SPECIFIC REQUIREMENTS

Although the following design parameters are considered site-specific, and are not quantified for this study, allowances for normal conditions and construction are included in the cost estimates.

- Flood plain considerations
- Existing soil/site conditions

- **Water discharges and reuse**
- **Rainfall/snowfall criteria**
- **Seismic design**
- **Buildings/enclosures**
- **Fire protection**
- **Local code height requirements**
- **Noise-regulations– Impact on site and surrounding area**

2.7 FUEL PROPERTIES

2.7.1 Coal Characteristics

Design coal feed to the CTL plants is Illinois #6 bituminous coal with characteristics presented in Table 2-3. The as-received coal properties shown in Table 2-3 are from the QGESS Detailed Coal Specifications document. The as-received coal is dried to 5% moisture before being pressurized and fed to the gasifier.

Table 2-2
Illinois #6 Coal Specification

Rank	HV Bituminous	
Seam	Illinois #6 (Herrin)	
Source	Franklin Co., IL	
Ultimate Analysis, weight%	As-Received	Dry
Carbon	63.75	71.72
Hydrogen	4.50	5.06
Nitrogen	1.25	1.41
Chlorine	0.29	0.33
Sulfur	2.51	2.82
Oxygen	6.88	7.75
Ash	9.70	10.91
Moisture	11.12	0.00
Total	100.0	100.0
Proximate Analysis, weight%	As-Received	Dry
Volatile Matter	34.99	39.37
Fixed Carbon	44.19	49.72
Ash	9.70	10.91
Moisture	11.12	0.00
Total	100.0	100.0
Higher Heating Value (HHV), Btu/lb	11,666	13,126
Sulfur Analysis*, weight%		Dry
Pyritic		1.14
Sulfate		0.22
Organic		1.46
Mercury, ppmW (moisture-free basis)		0.150
Ash Fusion Temperatures at Reducing Conditions, °F		
Initial Deformation	2,194	

Softening	2,260
Hemispherical	2,345
Fluid	2,415

*In accordance with NETL CTL Study, this study assumes that all sulfur in the coal is converted in the gasifier and leaves with the syngas

2.8 ENVIRONMENTAL DESIGN TARGETS

The environmental limits presented in this section refer to the gasification/power cycle only, because the environmental requirements for the FT plant are considered beyond the scope of the study.

The environmental targets for the study were considered on a technology- and fuel-specific basis. Since all the cases are located at a greenfield site, permitting a new plant would involve the New Source Review (NSR) permitting process. The NSR process requires installation of emission control technology, meeting either the best available control technology (BACT) determinations for new sources located in areas meeting ambient air quality standards, or the lowest achievable emission rate (LAER) technology for sources located in areas that do not meet ambient air quality standards. This study uses the BACT guidelines, summarized in Table 2-7.

Table 2-3
BACT Environmental Design Basis

Pollutant	Control Technology	Limit
Sulfur Oxides (SO ₂)	Rectisol® + Claus Plant/	≤ 0.050 lb/MMBtu
Nitrogen Oxides (NO _x)	Low NO _x Burner (LNB) and N ₂ Dilution primarily with humidification as needed	15 ppmvd (@ 15% O ₂)
Particulate Matter (PM)	Cyclone/Barrier Filter/Wet Scrubber/AGR Absorber	0.006 lb/MMBtu
Mercury (Hg)	Activated Carbon Bed	95% removal

Total air pollutants in all vents must meet the above specifications even if atmospheric venting is minimal.

2.9 CO₂ PRODUCT TREATING AND PURIFICATION DESIGN CRITERIA

For this study, recovered CO₂ is delivered at the battery limit (B/L), with specifications listed in Table 2-5 for saline reservoir sequestration per the QGESS CO₂ Impurities Design Parameters document.

Table 2-4
CO₂ Product Specifications

Compositions:		Conceptual Design	Range in Literature
CO ₂ ,	vol% (Min)	95	90 - 99.8
N ₂	vol% (Max)	} 4	0.01 - 7
Ar	vol% (Max)		0.01 - 4
O ₂ ,	vol% (Max)		0.01 - 4
H ₂	vol% (Max)		0.01 - 4
CH ₄ ,	vol% (Max)		0.01 - 4
H ₂ O,	ppmv (Max)	300	20 - 650
H ₂ S	vol% (Max)	0.01	0.002 - 1.3
NH ₃	ppmv (Max)	50	0 - 50
SO ₂ ,	ppmv (Max)	100	10 - 50,000
NO _x ,	ppmv (Max)	100	20 - 2,500
CO,	ppmv (Max)	35	10 - 5,000
B/L pressure, psia		2,215	

CO₂ compression and drying facilities were provided to boost the CO₂ product pressure to the required B/L requirement

2.10 RAW WATER USAGE

Raw water makeup was assumed to be provided 50% by a publicly owned treatment works and 50% from groundwater. It was assumed that all cases utilized a mechanical draft, evaporative cooling tower, and all process blowdown streams were assumed to be treated and recycled to the cooling tower. The design ambient wet bulb temperature of 11°C (51.5°F) was used to achieve a cooling water temperature of 16°C (60°F) using an approach of 5°C (8.5°F). The cooling water range was assumed to be 11°C (20°F). The cooling tower makeup rate was determined using the following:

- **Evaporative losses of 0.8 percent of the circulating water flow rate per 10°F of range**
- **Drift losses of 0.001 percent of the circulating water flow rate**

Blowdown losses were calculated as follows: Blowdown Losses = Evaporative Losses / (Cycles of Concentration - 1) Where cycles of concentration is a measure of water quality, and a mid-range value of 4 was chosen for this study.

The water balances presented in subsequent sections include the water demand of the major water consumers within the process, the amount provided by internal recycle, the amount of raw water withdrawal by difference, the amount of process water returned to the source and the raw water consumption, again by difference.

2.11 POWER GENERATION & AUXILIARY LOADS

For the power generation portion in each of the three cases in the current study, RTI developed an Aspen model for the heat recovery steam generator and the steam turbines. This model takes into account process steam generated/consumed in the gasification and syngas cleanup island

The models were validated against data from Case B1B of the DOE baseline study on bituminous coal for IGCC applications [16]. This model was later extended to include the steam usage (generation/consumption) in the CTL island. Auxiliary loads were estimated, wherever applicable, by pro-rating from the DOE CTL study using relevant scaling parameters obtained from the model's heat and material balance. In cases where there was not sufficient level of details available regarding gasification island processes, the auxiliary loads were pro-rated using the Case B1B of the bituminous coal to IGCC study. For the Rectisol® and Product Upgrading processes, the data from the DOE sponsored Bechtel study were used.

The AR compact gasification system and RTI advanced syngas cleanup and AACRP systems have different auxiliary loads that are not scalable with some of the loads specified in the DOE CTL study or the DOE Bituminous coal to IGCC study. These loads were calculated using the RTI model for the RTI WDP and DSRP for RTI technologies. For AR-related technologies, the power loads were scaled using the data provided by AR for the IGCC and CTM studies conducted as part of this project.

2.12 CAPITAL COST ESTIMATION METHODOLOGY

2.12.1 General

For the CTL and GTL plant with CO₂ capture the DOE Baseline Studies provided a code of accounts grouped into 14 major systems. Each of these major systems is broken down further into different subsystems. This type of code-of-accounts structure has the advantage of grouping all reasonably allocable components of a system or process into a specific system account.

For the three CTL cases evaluated in this study, cost estimation for systems that are not related to the AR advanced gasification and RTI advanced syngas cleanup and AACRP systems follows the capital cost scaling guidelines and parameters described in the NETL Capital Cost Scaling Methodology document. In general, this cost estimation methodology involves determining the scaling parameters, exponents and coefficients from the Capital Cost Scaling Methodology, as well as the reference cost and baseline capacity from the CTL and GTL studies. Once these have been established, the capital cost can be estimated based on the revised capacity from the HMB developed by RTI's ASPEN models for the CTL cases.

Costs for equipment in the AR advanced gasification system that are proprietary to AR, such as its DSP and compact gasifier, were scaled with similar costs for the IGCC and CTM cases provided by AR. For auxiliary equipment within the AR gasification section that is not included in the technology licensor's package, costs were scaled up with similar costs from the IGCC and CTM studies developed during this project.

RTI performed a bottoms-up, major equipment factored cost estimation for the RTI advanced syngas cleanup and AACRP systems. These are described in greater detail in Section 2.12.2.

2.12.2 RTI WDP System Capital Cost Estimate Criteria

For WDP, RTI's process model incorporates actual data from the pilot plant testing at Eastman Chemical and the 50 MW demonstration plant at Tampa Electric Company. For AACRP, RTI developed a detailed process model using ProMax with TSWEET kinetics for the CO₂-amine reactions. This process model was reviewed by a recognized expert in the field of acid gas

removal technologies. RTI has diligently worked to validate these models against actual pilot and demonstration plant data for RTI's proprietary technology and any publicly available DOE reports for the more commercial technologies.

Using these models, RTI developed HMBs and sized equipment lists. Using process conditions and experience gained from the pilot plant and demonstration plant operation, materials of construction were identified. With the sized equipment list and materials of construction, RTI developed equipment cost estimates from actual vendor quotes or scaled equipment costs from the 50 MW demonstration unit for a majority of the equipment. For some of the more generic and non-proprietary equipment, the equipment costs were developed using Aspen Plus ICARUS Process Economic Analyzer. RTI then used the same factored estimation methods used in the DOE CTL Study to estimate bare erected costs (BEC) for WDP, WGS and AACRP. This BEC information was used directly in its calculation of the overall CTL plant capital expenditure (CAPEX). In addition to using the actual pilot plant and demonstration plant data for developing equipment costs, RTI used this available data to develop utility and variable cost especially for sorbent, WGS catalysts, and CO₂ capture solvent.

2.12.3 Fischer-Tropsch Synthesis and Product Upgrade

The DOE sponsored Bechtel GTL study provides extensive details on the FT-synthesis and Product Upgrade process flow and plant layout that are not available from the more recent DOE CTL and GTL studies. The two GTL studies, Bechtel and DOE, are similar in their approach but the DOE study lacks the process details that are available in the Bechtel study. Hence, the Bechtel GTL study was used as a basis to model the FT synthesis and product upgrading blocks for the three comparison cases. The models were used to determine the scaling parameters for the FT synthesis and Product Upgrading blocks. The exponents for scaling were taken from the Bechtel study. Using the scaling parameters and the exponents determined, the costs for the equipment were scaled using costs from the DOE GTL study for the corresponding block. The DOE GTL study was used for the capital cost scaling since the process flow is similar between the Bechtel and the DOE study, and the DOE study, being more recent, is likely to have more updated costs. The GTL study was chosen over the CTL study since the GTL process uses the cobalt catalyst with a H₂/CO ratio of 2, and is similar to the process used in this work. The scaling parameters and exponents derived from the Bechtel study and used in this work are shown in Table 2.5.

Table 2.5
Scaling Parameters and Exponents used for FT synthesis and Upgrading

Plant	Reference stream	Units	Exp
FT-synthesis	gas feed rate	MM SCFH	0.75
CO ₂ removal	syngas feed rate	MM SCFH	0.76
CO ₂ removal	CO ₂ stream	MM SCFH	0.76
Dehydration/Compression	inlet feed rate	MM SCFH	0.75
Hydrocarbon recovery	syngas feed	MM SCFH	0.65
H ₂ recovery	product H ₂	MM SCFH	0.55
Wax hydrocracking	wax feedrate	MIbs/hr	0.55
Distillate hydrotreating	liquid feed rate	MIbs/hr	0.60

Naphtha hydrotreating	C7+ feedrate	MIbs/hr	0.65

2.12.4 Balance of Plant Capital Cost Estimate Criteria

The capital cost estimates for the rest of the CTL systems that are unrelated to coal gasification, sulfur and CO₂ removal, and FT synthesis and product upgrade were developed based on the DOE CTL study. The costs were adjusted for differences in unit or plant capacity per NETL's Guidelines as described in the NETL Capital Cost Scaling Methodology document

Table 2-8 shows the code of accounts for the CTL plant. These systems are further broken down to include the various subsystems. The scaling parameters for these balance-of-plant (BOP) processes, as laid out by the NETL Capital Cost Scaling Methodology document, are also shown in this table.

Table 2-5
Code of Accounts for Report CTL Plant

Acct No.	Item/Description	Scaling Parameter
1	COAL & SORBENT HANDLING	
1.1	Coal Receive & Unload	Coal Feed Rate
1.2	Coal Stackout & Reclaim	Coal Feed Rate
1.3	Coal Conveyors & Yard Crush	Coal Feed Rate
1.4	Other Coal Handling	Coal Feed Rate
1.9	Coal & Sorbent Handling Foundations	Coal Feed Rate
2	COAL & SORBENT PREP & FEED	
2.1	Coal Crushing & Drying	Coal Feed Rate
2.2	Prepared Coal Storage & Feed	Coal Feed Rate
2.3	Dry Coal Injection System	Coal Feed Rate
2.4	Misc Coal Prep & Feed	Coal Feed Rate
2.9	Coal & Sorbent Feed Foundation	Coal Feed Rate
3	FEEDWATER & MISC. BOP SYSTEMS	
3.1	Feedwater System	BFW (HP only)
3.2	Water Makeup & Pretreating	Raw Water Makeup
3.3	Other Feedwater Subsystems	BFW (HP only)
3.4	Service Water Systems	Raw Water Makeup
3.5	Other Boiler Plant Systems	Raw Water Makeup
3.6	FO Supply Sys and Nat Gas, incl Pipeline	Coal Feed Rate
3.7	Waste Treatment Equipment	Raw Water Makeup
3.8	Misc Power Plant Equipment	Gross Power Output
4	GASIFIER & ACCESSORIES	
4.1	Gasifier, Syngas Cooler & Auxiliaries	Syngas Throughput
4.3	ASU/Oxidant Compression	O ₂ Production
4.4	LT Heat Recovery & FG Saturation	Syngas Flow
4.6	Flare Stack System	Syngas Flow
4.9	Gasification Foundations	Syngas Flow
5A	GAS CLEANUP & PIPING	
5A.1	Rectisol [®] System	Calculated
5A.2	Elemental Sulfur Plant	Calculated
5A.3	Mercury Removal	Estimated Hg Flow
5A.4	Shift Reactors	Calculated
5A.6	Blowback Gas System	Syngas Flow
5A.7	Fuel Gas Piping	Fuel Gas Flow
5A.9	HGCU Foundations	Sulfur Production
5AA	FT SYNTHESIS AND PRODUCT UPGRADE	
5AA.1	FT Synthesis	Syngas Feed Rate
5AA.2	Amine CO ₂ Removal	CO ₂ removed
5AA.4	Compression	Compressor feed
5AA.5	Hydrocarbon Recovery	Syngas feed
5AA.6	Hydrogen Recovery	Product H ₂
5AA.8	Naphtha Hydrotreater	C ₇₊ Feed Rate
5AA.9	Diesel Hydrotreater	Liquid Feed Rate
5AA.10	Wax Hydrocracker	Wax Feed Rate
5AA.11	Raw Fuel Gas Compressor, HP	Fuel Gas
5AA.12	Raw Gas to GT Compressor, HP	Fuel Gas
5B	CO ₂ REMOVAL & COMPRESSION	
5B.1	CO ₂ Removal System	Calculated

Acct No.	Item/Description	Scaling Parameter
5B.2	CO ₂ Compression & Drying	CO ₂ Product
6	COMBUSTION TURBINE/ACCESSORIES	
6.1	Combustion Turbine Generator	Fuel Gas Flow
6.9	Combustion Turbine Foundations	Fuel Gas Flow
7	HRSG, DUCTING & STACK	
7.1	Heat Recovery Steam Generator	HRSG Duty
7.3	Ductwork	Vol Flow to Stack
7.4	Stack	Vol Flow to Stack
7.9	HRSG, Duct & Stack Foundations	Vol Flow to Stack
8	STEAM TURBINE GENERATOR	
8.1	Steam TG & Accessories	Turbine Capacity
8.2	Turbine Plant Auxiliaries	Turbine Capacity
8.3a	Condenser & Auxiliaries	Condenser Duty
8.3b	Air Cooled Condenser	Condenser Duty
8.4	Steam Piping	BFW (HP Only)
8.9	TG Foundations	Turbine Capacity
9	COOLING WATER SYSTEM	
9.1	Cooling Towers	Cooling Tower Duty
9.2	Circulating Water Pumps	Circ H ₂ O Flow Rate
9.3	Circ. Water System Auxiliaries	Circ H ₂ O Flow Rate
9.4	Circ Water Piping	Circ H ₂ O Flow Rate
9.5	Makeup Water System	Raw Water Makeup
9.6	Component Cooling Water System	Circ H ₂ O Flow Rate
9.9	Circ. Water System Foundations	Circ H ₂ O Flow Rate
10	ASH/SPENT SORBENT HANDLING SYS	
10.1	Slag Dewatering & Cooling	Slag Production
10.6	Ash Storage Silos	Slag Production
10.7	Ash Transport & Feed Equipment	Slag Production
10.8	Misc. Ash Handling System	Slag Production
10.9	Ash/Spent Sorbent Foundation	Slag Production
11	ACCESSORY ELECTRIC PLANT	
11.1	Generator Equipment	Turbine Capacity
11.2	Station Service Equipment	Auxiliary Load
11.3	Switchgear & Motor Control	Auxiliary Load
11.4	Conduit & Cable Tray	Auxiliary Load
11.5	Wire & Cable	Auxiliary Load
11.6	Protective Equipment	Auxiliary Load
11.7	Standby Equipment	Total Gross Output
11.8	Main Power Transformers	Total Gross Output
11.9	Electrical Foundations	Total Gross Output
12	INSTRUMENTATION & CONTROL	
12.4	Other Major Component Control	Auxiliary Load
12.6	Control Boards, Panels & Racks	Auxiliary Load
12.7	Computer & Accessories	Auxiliary Load
12.8	Instrument Wiring & Tubing	Auxiliary Load
12.9	Other I & C Equipment	Auxiliary Load
13	IMPROVEMENT TO SITE	
13.1	Site Preparation	Accounts 1-12
13.2	Site Improvements	Accounts 1-12
13.3	Site Facilities	Accounts 1-12

Acct No.	Item/Description	Scaling Parameter
14	BUILDING & STRUCTURES	
14.1	Combustion Turbine Area	Gas Turbine Power
14.2	Steam Turbine Building	Accounts 1-12
14.3	Administration Building	Accounts 1-12
14.4	Circulation Water Pump House	Circ H ₂ O Flow Rate
14.5	Water Treatment Buildings	Raw Water Makeup
14.6	Machine Shop	Accounts 1-12
14.7	Warehouse	Accounts 1-12
14.8	Other Buildings & Structures	Accounts 1-12
14.9	Waste Treating Building & Structures	Raw Water Makeup

As Table 2-7 is based on DOE CTL study, it does not necessarily have account and/or subaccount numbers for the advanced technologies being evaluated in this study. To support optimal comparison of these advanced technologies with the existing commercial technologies, the advanced technologies used the same account and/or subaccount numbers as the existing commercial technologies that they are most analogous to. For technologies without a defined account number, one was created. Table 2-8 provides a list of the advanced technologies evaluated in this study and their associated account numbers.

Table 2-6
Code of Accounts for Advanced Technologies being Evaluated

Account Number	Title	Advanced Technology
2.3	Dry Coal Injection System	AR DSP
4.1	Gasifier, Syngas Cooler & Auxiliaries	AR Compact Gasifier
5A.1	AACRP	AACRP
5A.2	RTI DSRP	RTI DSRP
5A.4a	RTI WDP System	RTI WDP
5A.4b	Shift Reactors	AFWGS

2.12.5 Home Office, Engineering Fees and Project/Process Contingencies

Engineering and construction management fees and home office cost, project and process contingencies were factored from each subsystem's BEC. These were then added to the BEC to come up with the TPC of the system.

Process contingency is intended to compensate for uncertainty in cost estimates caused by performance uncertainties associated with the development status of a technology, Table 2-9 shows the process contingencies applied for the technologies of interest in this study.

Table 2-7
Process Contingencies for Advanced Technologies being Evaluated

Account Number	Advanced Technology	Process Contingency (% of Associated Process Capital)	Rationale
2.3	AR DSP	20%	The AR DSP and static splitter feed system is currently demonstrated at the small pilot level. The system shares similar types of commercially demonstrated equipment with the conventional lockhopper feed system e.g. feed bins, compressors.
4.1	AR Compact Gasifier	25%	At 25%, this is within range of contingencies applied to technologies demonstrated at the small pilot level
5A.1	AACRP	20%	This is consistent with the contingency applied for similar processes such as Selexol and Rectisol [®] in the DOE/NETL studies.
5A.2	RTI DSRP	20%	The DSRP technology has been demonstrated in RTI's pilot testing at Eastman Chemical Company (DE-AC26-99FT40675)
5A.4a	RTI WDP	20%	The WDP technology has been demonstrated in RTI's 50 MW demonstration plant project (DE-FE0000489)
5A.4b	AFWGS	0%	Sweet water gas shift technology is commercially proven

2.12.6 Owner's Cost

Owner's cost was then added to TPC to come up with the total overnight cost (TOC) for the system. Owner's costs as defined in the DOE CTL Study include the following:

- Preproduction Costs –
 - 6 months of all labor cost
 - 1 month of maintenance materials
 - 1 month of non-fuel consumables
 - 1 month of waste disposal
 - 25% of 1 month fuel cost at 100% capacity factor
 - 2% TPC
- Inventory Capital -
 - 60-day supply of fuel and consumable at 100% CF
 - 0.5% TPC (spare parts)
- Initial Cost for Catalyst and Chemicals per design
- Land Cost = \$900,000 for 300 acres x \$3,000/acre
- Other Owner's Costs at 15% TPC
- Financing Costs at 2.7% TPC

2.13 OPERATION & MAINTENANCE COSTS

The operation and maintenance (O&M) costs pertain to those charges associated with operating and maintaining the CTM plants over their expected life. These costs include:

- Operating labor
- Maintenance – material and labor
- Administrative and support labor
- Consumables
- Fuel
- Waste disposal

There are two components of O&M costs; fixed O&M, which is independent of FT liquids and power generation, and variable O&M, which is proportional to FT liquids and power generation. Variable O&M costs were estimated based on 90% capacity factor.

2.13.1 Fixed Costs

Operating labor cost was determined based on the number of operators required to work in the plant. Other assumptions used in calculating the total fixed cost include:

• 2011 Base hourly labor rate, \$/hr	\$39.7
• Length of work week, hrs	50
• Labor burden, %	30
• Administrative/Support labor, % O&M Labor	25
• Maintenance material + labor, % TPC	2.8
• Maintenance labor only, % maintenance material + labor	35
• Property Taxes and insurances, % TPC	2

2.13.2 Variable Costs

The cost of consumables, including fuel, was determined based on the individual rates of consumption, the unit cost of each specific consumable commodity, and the plant annual operating hours. Waste quantities and disposal costs were evaluated similarly to the consumables.

The unit costs for major consumables and waste disposal was selected from the DOE/NETL 1477 Baseline Report, QGESS Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases [15] and from the QGESS Fuel Prices for Selected Feedstocks in NETL Studies [8] document.

The 2011 coal price as delivered to the Midwestern, USA CFTL plant is \$36.57/ton, per the QGESS Fuel Prices for Selected Feedstocks in NETL Studies document

2.13.3 CO₂ Transport and Storage Costs

CO₂ Transport and Storage (T&S) costs used for the Midwestern plant location is \$11/tonne. The required selling prices for naphtha and diesel are reported both with and without the cost of CO₂ T&S.

2.14 FINANCIAL MODELING BASIS

2.14.1 Required Selling Price

The figure-of-merit used to evaluate the CFTL plant's overall financial performance is the estimated crude oil equivalent required selling price (RSP) of FTL products, expressed in \$/bbl of product. All costs are expressed in the "first-year-of-construction" year dollars, and the resulting RSP is also expressed in "first-year-of-construction" year (2011) dollars.

Two financial structures were assumed for calculating the figures-of-merit: (i) a financial structure representative of a commercial fuels project, and (ii) a financial structure with loan guarantees or other government subsidies. The financial assumptions and structures used to estimate the RSPs are shown in Table 2-10 and Table 2-11.

Table 2-8
Financial Assumptions for RSP Calculation

Parameter	Value
TAXES	
Income Tax Rate	38% (Effective, 34% Federal, 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	None
Tax Holiday	None
FINANCING TERMS	
Repayment Term of Debt	30 years
Grace Period on Debt Repayment	None
Debt Reserve Fund	None
TREATMENT OF CAPITAL COSTS	
Capital Cost Escalation During Construction (nominal annual rate)	3.6%
Distribution of Total Overnight Capital over the Capital Expenditure Period (before escalation)	5 Year Period: 10%, 30%, 25%, 20%, 15%
Working Capital	Zero for all parameters
% of Total Overnight Capital that is Depreciated	100%
INFLATION	
RSP, O&M, Fuel Escalation (nominal annual rate)	3.0% RSP, O&M, COE, Fuel

Table 2-9
Financial structure for FT CTL projects

SCENARIO	COMMERCIAL FUELS	LOAN GUARANTEES
% Debt	50%	60%
% Equity	50%	40%
Debt Interest Rate	8% (LIBOR = 3.5% + 4.5%)	4.56% (CMT = 4.34% + 0.22%)
Internal Rate of Return on Equity	20%	20%
After Tax Weighted Cost of Capital	12.48%	9.70%
Capital Charge Factor (CCF)	0.218	0.170

The same financial modding methodology was used for this study as per the DOE CTM& CTL Studies. This is a simplified method that is a function of the plant TPC, capital charge factor, fixed and variable operating costs, capacity factor and net power generation, as shown in the equation below:

$$\text{RSP} = \frac{\text{first year capital charge} + \text{first year fixed operating costs} + \text{first year variable operating costs}}{\text{Annual net Fischer-Tropsch Liquids Production}}$$

$$RSP = \frac{(CCF)(TOC) + OC_{fix} + (CF)(OC_{var})}{\text{Annual net Fischer-Tropsch Liquids Production}}$$

The capital charge factors (CCF) used in evaluating the RSPs were pre-calculated using the NETL Power Systems Financial Model (PSFM). These factors are valid for global economic assumptions used for the two pre-determined finance structures and capital expenditure periods. The resulting CCFs for the commercial fuels and loan guarantee finance structures are 0.218 and 0.170 respectively.

The petroleum equivalent diesel price and the equivalent crude oil price are calculated as per the DOE CTL study [2]. The petroleum equivalent diesel price is calculated by taking the cost of production for diesel in \$/bbl and multiplying this value by the ratio of the lower heating values (LHV) of FT diesel and petroleum diesel.

$$\text{Petroleum Equivalent Diesel Price} = \left(\frac{\text{Petroleum Diesel}}{LHV_{FT Diesel}} \right) * COP_{FT Diesel}$$

The equivalent crude oil price is then calculated by multiplying the petroleum equivalent diesel price by a factor of 0.80. The factor of 0.80 was calculated from data of historic spot prices provided by the EIA from June 2009 through November 2013 for various fuel types.

$$\text{Equivalent Crude Oil} = 0.80 * \text{Petroleum Equivalent Diesel Price}$$

2.14.1 Cost of CO₂ Captured/Avoided

The cost of CO₂ capture requires data for a non-capture plant. Since the CO₂ removal in a CTL plant is carried out for process requirements, the difference between a CTL with CO₂ capture and a non-CO₂ capture plant is the compression of the removed CO₂ to 2215 psia for sequestration. The cost of the non-capture CTL plant was calculated by removing the costs for CO₂ capture from the analysis.

2.14.1 CO₂ Sales Price

Sensitivity analysis was performed to evaluate the impact of CO₂ sales price on the RSP of the FT-liquids. The varying parameter is the CO₂ sales price at the plant gate and ranges from \$0/tonne (baseline case assuming no value to the product CO₂) and \$60/tonne.

The formula used to calculate the revised RSP after considering the CO₂ sales is shown below:

$$RSP_{CO_2 \text{ Sales}} = \text{Baseline RSP} - \frac{(\text{CO}_2 \text{ Sales Price}) \times \text{annual tonnes of CO}_2 \text{ product}}{\text{Annual net Fischer-Tropsch Liquids Production}}$$

2.14.2 Cost of CO₂ Emissions

Sensitivity analysis on cost of CO₂ emissions on the RSP of the FT-liquids was also performed. The range of the CO₂ emissions cost was varied from \$0/tonne (baseline case assuming no CO₂ emissions cost) and \$60/tonne.

The formula used to calculate the revised RSP after considering the CO₂ emissions cost is shown below:

$$\text{RSP}_{\text{CO}_2 \text{ Emissions}} = \text{Baseline RSP} + \frac{(\text{Cost of CO}_2 \text{ Emissions}) \times \text{annual tonnes of CO}_2 \text{ emitted}}{\text{Annual net Fischer-Tropsch Liquids Production}}$$

3.1 PROCESS OVERVIEW

The conversion of coal-to-liquids via FT Synthesis is a two-step process: first conversion of the coal to the appropriate quality syngas via gasification followed by the water-gas shift reaction, and second, conversion of the syngas to FT liquids by catalytic conversion and subsequent product upgrade to FT-naphtha and FT-diesel.

Syngas is generated from the gasification of Illinois #6 coal in a high-pressure, oxygen-blown gasifier. For the syngas to be usable in the FT synthesis reactor, the syngas needs to be shifted to increase the hydrogen content so that a H_2/CO molar ratio of $\sim 2:1$ in the feed stream to the FT-synthesis reactor can be achieved. This can be carried out using either a sour water-gas shift (SWGS) reaction where the shift occurs prior to sulfur removal using sulfur-tolerant WGS catalyst, or the high temperature WGS which occurs after the sulfur removal step. Sour water gas shift step complements the Rectisol[®] process by avoiding the need to re-heat the syngas for the sweet water shift process after the sulfur removal step. The syngas also needs to be cleaned of particulate matter, mercury, sulfur species (H_2S , COS), and HCl. Sulfur and CO_2 can be removed from the syngas via the dual-stage Rectisol[®] process or using a combination of the RTI WDP and CO_2 capture processes such as activated MDEA solvents. Rectisol[®] produces an acid gas stream consisting of CO_2 and H_2S and elemental sulfur is recovered using the Claus process. In the case of RTI WDP, sulfur is removed from the syngas as an SO_2 stream. This can be used to produce elemental sulfur using the RTI DSRP or a modified-Claus process where a portion of the SO_2 stream is converted to H_2S stream to provide a $2:1 H_2S:SO_2$ stream required as feed for the Claus process. Both processes can reduce sulfur content in the syngas to below 0.1 ppm, required for the FT-synthesis process. A separate COS hydrolysis step is not needed as both Rectisol[®] and RTI WDP process remove both H_2S and COS. HCl is either removed using a water-wash scrubber or using a chloride guard bed when coupled with the RTI warm gas cleanup system (note that such a guard bed is only required for WDP because of the high level of HCl found in syngas produced from Illinois #6 coal and would not normally be needed for most other coals). Since the RTI WDP process removes sulfur at warm gas temperatures, either a sour or sweet water-gas shift can be used. The syngas is cooled in a low-temperature heat recovery system and then cleaned of mercury in a fixed-bed reactor. A ZnO bed is used as a sulfur guard bed upstream of the FT-synthesis process to prevent catalyst deactivation, in the event of an upset in the sulfur removal step. The treated syngas is expanded in a turboexpander to the pressures required for FT synthesis and is fed to the FT-synthesis reactors to generate FT-liquids.

A low-temperature FT (LTFT) process is used in this study. A Co-based catalyst process was chosen for this study over the Fe-based catalyst system to allow the CO_2 capture process to occur upstream of the FT-synthesis process. This enables the CO_2 capture to be carried out more efficiently at higher pressures closer to the gasifier pressures compared to the lower pressures encountered after the FT-synthesis reactors. After the FT synthesis, the FT-products are further upgraded to produce a 30:70 split of naphtha and diesel.

The tail gas from the FT process is used for coal drying and as fuel gas in the FT process. The remaining fuel gas is combusted in a gas turbine to generate power. The exhaust gas is used in a HRSG to recover heat and produce steam that is combined with the steam from the gasification and the FT areas to produce additional power in a steam turbine. The power produced in the combined cycles is used to supply the auxiliary load for the rest of the plant including gasifier for syngas production and cleanup and in the FT synthesis and upgrade processes. Any excess power is exported to the grid.

3.2 CTL PLANT COMMON PROCESS AREAS

As shown in Figure 2-1 in Section 2.2.1, the Case 5A reference plant consists of the following major process and/or utility blocks. Some of these blocks, or process areas, are common to the Case 5B and 5C plant configurations. These common process areas are in bold and italicized.

- Coal Sizing Handling
- Coal Prep, Drying
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- **Dry Coal Feed & Shell SCGP Gasifier System**
- **Syngas Cooling (Quench, Scrubbing Steam Generation)**
- **Gas Cleaning (Filters, WGS & AGR)**
- Mercury Removal
- CO₂ Compression and Purification Facilities
- **Sulfur Plant**
- Syngas Conditioning
- FT-Synthesis and Product Upgrade
- HRSG, Ducting and Stack
- Cooling Water Systems
- BFW/Condensate System
- Slag Recovery and Handling
- Accessory Electric Plant (Steam and Gas Turbines)
- Instrumentation and Control

The common areas are presented in brief here for general background information, and to avoid unnecessary repetition while discussing the individual cases. Detailed descriptions of the process related to the syngas generation and power production areas can be found in Section 3 of the DOE Baseline Study for Bituminous Coal (IGCC) to Electricity [16]. Detailed description of the FT-synthesis and product upgrade process can be found in the DOE-sponsored Bechtel study on GTL [4]. Where there is case specific performance information, these features are presented in the relevant case sections.

3.2.1 Synthesis Gas Production

3.2.1.1 Coal Milling, Grinding and Drying

The Shell process uses a dry feed system that is sensitive to the coal moisture content. For coal to flow smoothly through the lockhoppers that pressurize coal to the gasifier, the coal's surface moisture must be removed. The Illinois #6 coal used in this study contains 11.12% total moisture on an as-received basis. It is dried to 5% moisture for smooth flow through the dry feed system. The coal is simultaneously crushed and dried in the coal mill, then delivered to a surge hopper.

The coal is delivered to the site by 100-ton rail cars. It is unloaded into two receiving hoppers and fed to the vibratory feeder. It is then transferred through intermediate hoppers and silos to the coal crusher where it is reduced to 1-1/4" x 0 size.

3.2.1.2 Coal Preparation and Drying

The Shell process uses a dry feed system that is sensitive to coal moisture content. It was assumed that the coal must be dried to 5% moisture to allow for smooth flow through the dry feed system before feeding into the Shell entrained-flow gasifier. This moisture content is considered compatible with the storage, transport and feed injection requirements for the Shell entrained-flow gasifier and the Aerojet gasifier.

The drying heat is provided by burning tail-gas from the FT-plant in an incinerator with air as the oxidant.

3.2.1.3 Air Separation Unit

All cases include an ASU for generating oxygen. The ASU is a conventional, cryogenic, pumped liquid oxygen unit that provides oxygen for the gasification process, as well as nitrogen for ancillary equipment. The ASU is designed to produce 95 mol% O₂ for use in the gasifier and Claus plant (in Reference Plant Case 5A). The air compressor is powered by an electric motor. Nitrogen is recovered and used as a diluent for coal drying, as described earlier in 3.2.2. A small, separate stream of nitrogen is compressed to be used as a stripping gas in the Rectisol[®] process (in the Reference Plant Case 5A)

The battery limit conditions for the ASU products are summarized below:

Table 3-1
ASU Product Conditions

ASU Product	Pressure, psia	Temperature, °F
95% O ₂	23.2	55
Diluent N ₂	14.7	63
ASU Vent	16.4	64

3.2.1.4 Mercury Removal

Mercury removal is achieved by a packed bed of sulfur-impregnated activated carbon operating at 95°F. This packed-bed vessel is located upstream of the AGR unit with 20-second superficial gas residence time to achieve more than 90% removal of mercury in addition to removal of some portion of other volatile heavy metals such as arsenic. Mercury-removal systems using sulfur-impregnated activated carbon downstream of a coal gasifier have a reported bed life of 18 to 24 months, and usually replacement is required due to fouling of the bed rather than mercury saturation.

3.2.1.5 CO₂ Compression and Purification Facilities

CO₂ captured in the AGR process prior to the FT process and that captured in the amine block after the FT process is compressed to 2,215 psia by a multi-stage, intercooled centrifugal compressor that is equipped with a thermal swing adsorptive dryer, which dehydrates the CO₂ stream to a dew point of -40°F. Some CO₂ is withdrawn from this stream to be used for transporting the coal to the gasifier. The virtually moisture-free supercritical CO₂ stream is delivered to the plant battery limit as sequestration ready.

3.2.1.6 Slag and Ash Handling

Slag material drains from the gasifier into a water bath at the bottom of the gasifier vessel. The slag-water slurry is transferred to a slag crusher where the slag is crushed into pea size fragments. The slurry containing 5 to 10% solids is then transferred to a dewatering bin through a lockhopper. The separated water is clarified and reused as makeup to the water scrubber. The dried slag is stored for disposal.

3.2.1.7 Syngas Conditioning

While the AGR process (Rectisol® and RTI WDP with AACRP systems) can reduce the sulfur levels to < 1 ppm required for the FT-synthesis process, a sulfur polishing guard bed is added just prior to the FT-synthesis process to guard against any upset in the AGR process that may result in poisoning the FT-catalyst. The cost of the guard beds is insignificant compared to the cost of the shutdown and lost production resulting from an accidental upset in the Rectisol® or the RTI WDP process.

There are two ZnO fixed-beds that operate at 600°F. Syngas from the AGR process (Rectisol® for Case 5A and AACRP for Cases 5B and 5C) is heated in a series of heat exchangers, first using low-pressure steam to ~180-200°F, followed by a gas-gas interchanger that exchanges heat with the syngas coming out of the ZnO beds and heats the syngas to 550°F. The syngas is further heated to the operating temperature of 600°F by exchanging heat using the shifted syngas from the WGS process.

The cleaned syngas coming out of the ZnO guard beds and after the syngas interchanger is at 230°F. The syngas is further expanded in a turbine that reduces the syngas pressure to 371 psia and cools the syngas to 200°F. This syngas is then fed to the FT-synthesis reactors.

3.2.2 Fischer-Tropsch Synthesis

The FT-synthesis process chosen for this study uses a cobalt-catalyzed, LTFT process. Figure 3-1 shows a simplified BFD of the FT plant and includes FT Synthesis, CO₂ Removal, Recycle Gas Compression and Dehydration, Hydrocarbon Recovery and Hydrogen Recovery.

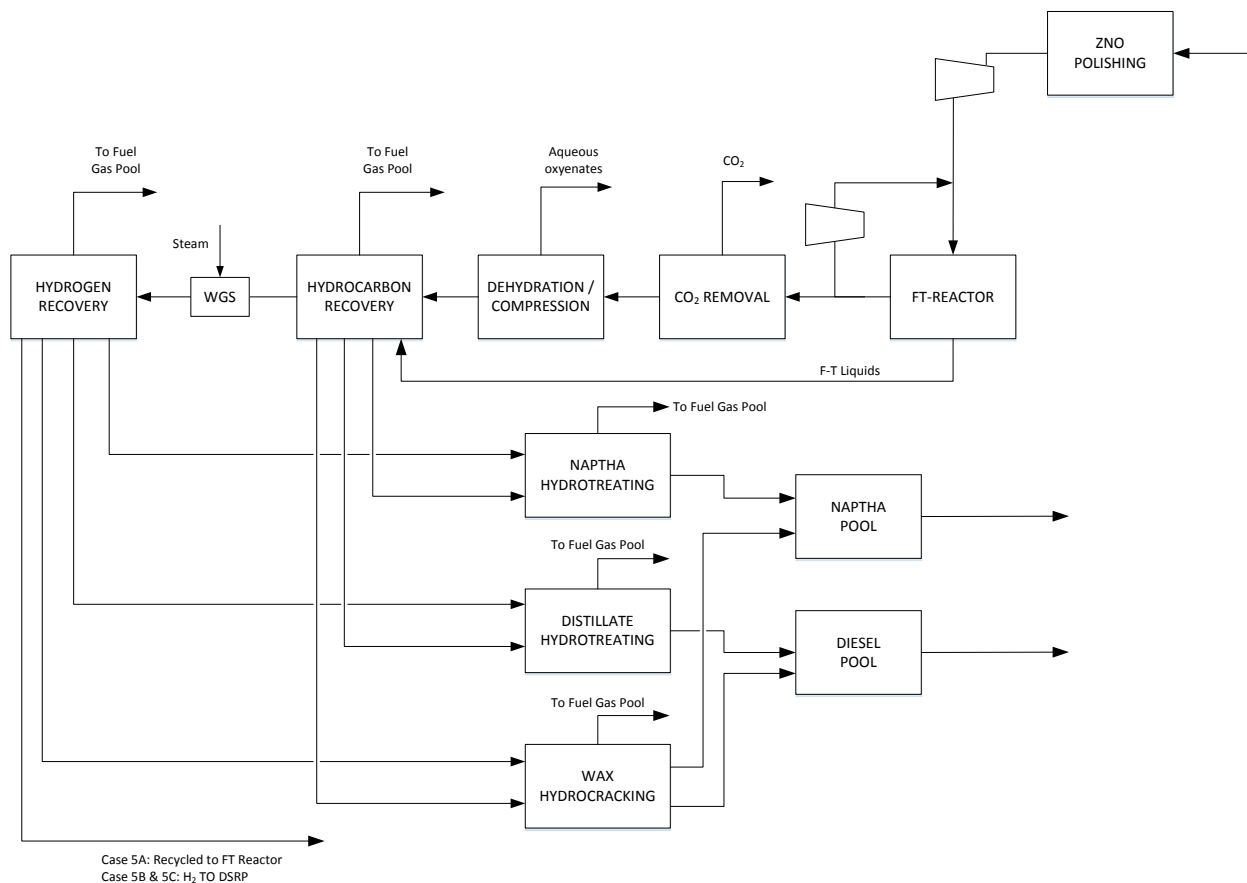


Figure 3-1 Simplified block flow diagram for the FT process.

3.2.2.1 FT Reactor

Each FT synthesis train consists of two parallel first-stage slurry-bed reactors that feed a common second-stage slurry-bed. The unconverted syngas leaving the first stage reactors is cooled to 150°F to condense and remove water and hydrocarbons before being re-heated and fed to the single-stage second reactor. The CO conversion in each slurry-bed reactor is assumed to be 61%, leading to a per pass conversion of 85%.

The FT reactors operate nominally at 428°F. The first-stage reactor operates at 335 psig and the second-stage reactor operates at 289 psig. Reactor temperature, which would increase due to the exothermic heat released by the FT reaction, is controlled by generating 190 psig steam through bayonet tubes suspended from an internal double tubesheet in the slurry-bed reactor. Although

the DOE CTL study also uses steam generation to control reactor temperature, these higher operating temperature results in 360 psig steam rather than 190 psig steam.

The vapor stream from the second-stage FT-reactor is passed through a cyclone separator to recover any entrained catalyst particles. It is then cooled to 100°F and sent to a three-phase separator to separate out an aqueous water stream, a liquid hydrocarbon stream and the vapor stream. The aqueous streams from the first and second stage reactors are combined and sent to the waste water treatment plant. The liquid hydrocarbon streams are sent to the hydrocarbon recovery plant and the vapor stream is sent to the vapor phase oxygenates water wash to separate vapor phase oxygenates from the unconverted syngas. The waste water stream from the water wash column is also sent to the waste water treatment plant. The unconverted syngas, after oxygenates removal, is sent to the CO₂ recovery plant.

Wax/catalyst streams are withdrawn from the FT-reactors and passed through hydroclones to produce an overflow stream and an underflow stream. The underflow stream, containing roughly 56% of the catalyst is pumped back to the reactors. The overflow stream is passed through a valve to reduce its pressure to 70 psig and sent to a product separator. The pressure reduction results in separation of the dissolved gases from the liquid product stream. The released gases are cooled to 100°F to recover additional liquid. The vapors are used as fuel gas in the plant. The liquid is sent to an intermediate tank for additional catalyst separation. The recovered catalyst is sent back to the FT-reactors and the clear wax stream is sent to the hydrocarbon recovery plant.

The slurry in the FT reactors contain 28 wt% catalyst. Catalyst replacement is estimated at 0.1% per day of total catalyst inventory, resulting in a catalyst lifetime of 2.7 years. For comparison, the catalyst lifetime expectation for Fe-based catalyst is 0.78 years.

3.2.2.2 *Recycle Gas Loop*

A recycle stream from unconverted vapor stream from the FT-synthesis is compressed and recycled upstream to the first FT-reactor, raising the overall conversion of FT synthesis to 85%.

3.2.2.3 *CO₂ Removal*

The remaining vapor stream is sent for CO₂ removal followed by product recovery. CO₂ removal is needed to prevent CO₂ solidification during the hydrocarbon recovery process where the syngas is cooled to -140°F to recover the heavier hydrocarbons as liquids. Unconverted syngas leaving the CO₂ removal plant should have the CO₂ concentration less than 400ppm. The captured CO₂ is combined with the CO₂ from the Rectisol[®]/AACRP unit in the gasification area, and sent for compression.

3.2.2.4 *Product Recovery*

The unconverted syngas is sent for dehydration where molecular sieve type dryers are used to reduce the moisture content of the syngas to approximately 0.1 ppm to prevent hydrate formation in the downstream cryogenic hydrocarbon recovery unit. After dehydration, the syngas is cooled to -140°F to condense out higher hydrocarbons from the syngas. Vapor and liquid streams are

recovered in a separator and re-heated to 85°F in heat exchangers. The vapor stream is sent for hydrogen recovery and the liquid stream is combined with the FT-liquids for product upgrading.

The FT-liquids for product upgrading are collected in the feed drum of the product fractionator where they are separated into a low-pressure light hydrocarbons vapor stream and three liquid streams, naphtha, middle distillate and wax. The three liquid streams are defined by the following boiling range cut points:

Naphtha	C5 saturates to 350°F (177°C)
Middle Distillate	350°F - 650°F (177°C - 343°C)
Wax	Greater than 650°F (343°C)

Details on the fractionation column are provided in the Bechtel and DOE reports [2-4].

3.2.2.5 Product Upgrading

Naphtha Hydrotreaters

In the naphtha hydrotreaters, the light liquid hydrocarbon products are processed to produce stabilized FT-naphtha. This is blended with the naphtha from the hydrocrackers to produce the crude naphtha product.

Middle Distillate Hydrotreaters

The fixed-bed distillate hydrotreater catalytically hydrotreats the FT middle distillates into a quality acceptable as diesel fuel. The hydrotreaters removes oxygenates and reduces the olefinic and aromatic content of the hydrocarbons. The hydrotreater also produces a residual light C₄-vapor stream and a waste water stream that are sent to the fuel gas and wastewater treatment plant, respectively.

Wax Hydrocracking

The wax hydrocracker breaks down the wax into naphtha and middle distillate, as well as some residual C₄-vapor which is recovered as fuel gas. The middle distillate from the wax hydrocracker is sent for blending with the middle distillate range product from the distillate hydrotreating section to produce diesel. The naphtha stream from the wax hydrocracker is combined with the naphtha product from the naphtha hydrotreater to produce crude naphtha product.

3.2.3 Heat Recovery and Power Generation

The gasification process generates a large amount of heat that is recovered via steam generation, for use as process requirements or for power production. In addition, the FT synthesis processes is an exothermic process and the reactor temperature is regulated by generating steam. The unconverted fuel gas from the FT synthesis and product upgrade, and in excess of that used to provide process heat for coal drying and within the FT-plant, is sent to Gas Turbine (GT) to generate power. The FT plant uses 3 GE MS6001B gas turbines, each producing a nominal

45 MW power. The exhaust gas from the GT is at 1020°F and is used further to extract heat for steam generation in the heat recovery steam generator (HRSG).

The process steam is generated at three different pressure levels, high pressure (HP) steam at 1815 psia, intermediate pressure (IP) steam at 458 psia and low pressure (LP) steam at 65 psia. The GT exhaust is at 1020°F and can only superheat the HP steam to 970°F. The IP steam is also reheated to 970°F. During the RTI's simulation of the power cycle, it was determined that the gas turbine flue gas outlet temperature was not high enough to superheat/reheat all the steam raised in the gasification and CTL section for power generation in the steam turbine.

LP exhaust steam from the last ST stage is condensed by splitting 50/50 to a surface condenser and an air-cooled condenser to conserve cooling water. The condensers operate at 0.698 psia with a corresponding condensing temperature of 90 °F.

The condensates are collected and sent to a deaerator to remove dissolve gases and treated to provide BFW for the steam generators.

3.2.3.1 Cooling Water Systems

Exhaust LP steam from the steam turbine is split 50/50 to a surface condenser cooled with cooling water and to an air-cooled condenser using ambient air and forced convection. The major impact of utilizing this parallel cooling method is a significant reduction in water requirement when compared to a wet cooling system.

The circulating water system is a closed-cycle cooling water system that supplies cooling water to the surface condenser to condense one-half of the main turbine exhaust steam. The system also supplies cooling water to the AGR plant as required, and to the auxiliary cooling system. The auxiliary cooling system is a closed loop process that utilizes a higher quality water to remove heat from compressor intercoolers, oil coolers and other ancillary equipment and transfers that heat to the main circulating cooling water system in plate-and-frame heat exchangers. The heat transferred to the circulating water in the surface condenser and other applications is removed by a mechanical draft cooling tower.

The system consists of two 50%-capacity vertical circulating water pumps (CWP), a mechanical draft evaporative cooling tower, and CS cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell wood frame counter-flow mechanical draft cooling tower.

3.2.3.2 BFW/Condensate System

The function of the feedwater (FW) system is to pump the various FW streams from the deaerator storage tank in the HRSG to the respective steam drums. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow. The FW pumps are supplied with instrumentation to monitor and alarm on low oil pressure, or high bearing

temperature. FW pump suction pressure and temperature are also monitored. In addition, the suction of each BFP is equipped with a startup strainer.

3.2.4 Water Balance

Water required for the operation of the facility is obtained from a source such as a lake, river, or well. If the quality of the water is adequate, raw water is used directly as makeup to the cooling tower and the gasifier quench. To meet the rest of the plant's water needs, makeup must be treated first by filtration to create service-quality water. This quality of water serves as makeup to the plant's potable water, demineralized water, fire water, and service water systems. Higher quality boiler feedwater is treated by a typical reverse osmosis and electrode ionization package. Water rejected by the system is of an acceptable quality to be used as makeup to the cooling tower.

In addition to meeting the makeup water needs of the facility, water treatment systems must be capable of capturing and treating on-site waste streams. Wastewater created by the gasification process must pass through several pretreatment steps before being combined with other wastewater streams. Metals, ammonia, and suspended solids are removed from the stream using a clarifier and a biological treatment unit. Once processed, the wastewater can be combined with the cooling tower blowdown as well as other plant waste streams in a final clarifier. Dechlorination and pH adjustment are performed as needed at this step of the process to meet all local discharge regulations. Solids separated out in this process are dried by means of a filter press and taken away for offsite disposal.

4.1 CASE 5A PROCESS DESCRIPTION

Case 5A is the reference plant design used for the CFTL study and was selected as a combination of two DOE case studies, “Cost and Performance Baseline for Fossil Energy Plants. Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis, Revised October 15, 2014, DOE/NETL-2011/1477” (CTL Study) and the “Analysis of Natural Gas-to-Liquid Transportation Fuels via Fischer-Tropsch”, Revised September 13, 2013, DOE/NETL-2013/1597 (GTL Study). The overall block flow diagram of the reference plant, referred to as Case 5A, is shown in Figure 4-1, with the accompanying stream flows shown in Table 4-1. In the section below, we have limited our discussion to the processes that vary between the three studied cases (Case 5A (Reference Case), Case 5B and Case 5C).

Coal Feed System

The dried coal is drawn from the surge hoppers and fed through a pressurization lock hopper system to a dense phase pneumatic conveyor, which uses a stream of high pressure CO₂ (~800 psia) withdrawn from the CO₂ compressors to convey the coal to the gasifiers.

Shell Gasifier

Syngas is generated from the gasification of Illinois #6 coal in a high-pressure, oxygen-blown Shell quench-type gasifier. The high-temperature entrained-bed gasifier system includes high temperature-gas cooling via syngas recycle, slag handling and solids waste handling equipment. In these gasifiers, the coal feedstock reacts with oxygen, at 615 psia, in a reducing environment to produce principally hydrogen and carbon monoxide with a small amount of CO₂ and CH₄, maintaining the reactor temperature at 2600°F.

High-temperature heat recovery in each gasifier train is accomplished in three steps, including the syngas cooling through gasifier membrane wall, which maintains a protective ash layer over the membrane wall, product syngas cooling from the gasifier using a syngas recycle quench to lower the temperature below the ash melting point followed by a syngas cooler to cool the hot syngas stream and a partial water quench up to 750°F. Syngas cooling contributes to the production of HP steam for the water gas shift reactors and power generation in the steam cycle.

The solids are removed as both slag and ash. Liquid slag is solidified in a water bath and removed via a lock hopper system.

The design size used in this study requires the use of eight gasifiers that process 21,006 tpd of as-received coal. **Recently Shell is providing gasifiers with a 3200 tpd coal intake capacity [19]. These new provisions can help reduce the number of gasifiers from eight to seven and possibly reduce the capital cost.**

Dry Solids Removal and Wet Scrubbing

Ash carried over with the syngas is removed in a ceramic candle filter. The collected ash is also removed via a lock hopper system. The raw syngas exiting the ceramic particulate filter enters a quench scrubber that washes the syngas in a countercurrent flow in two packed beds, which removes essentially all traces of entrained particles. The scrubber removes any possibility of PM that might have escaped the candle filter either through leakage from the filter seals or any undetected candle breakage.

In addition to removing any remaining particulates, the scrubber serves a very important function of removal of chlorides. The bottoms from the scrubber are sent to the slag removal and handling system for processing. Sour water needs to be treated for the chlorides and metals before discharge.

Water Gas Shift

Coal-derived syngas from the wet scrubber enters the sour shift and cooling section. To achieve a 2:1 ratio of H_2 to CO in the final syngas to the FT synthesis reactors, approximately 55 to 60% of the coal-derived syngas is shifted, with the remainder bypassing the shift reactors. Two shift reactors in series are used to achieve the desired H_2 /CO composition in the syngas. The exothermic heat after the first-stage shift reactor is recovered by producing steam. The exothermic heat after the second-stage shift reactor is used to preheat the syngas entering the first-stage shift reactor.

The shifted syngas from the second-stage shift reactor outlet mixes with the bypass syngas and is then sent to low temperature gas cooling for further cooling before entering the downstream Rectisol[®] unit for sulfur and CO_2 removal.

Low Temperature Gas Cooling

Syngas is cooled in several steps to effectively recover heat to maximize efficiency. As the shifted syngas is cooled, IP and LP steam, process condensate and feed water are heated.

The sour water treatment unit combines the effluent from both syngas water scrubber and the process condensate from the low temperature syngas cooling and send them to a flash drum. Liquid from the flash drum bottom is preheated and fed to the stripper, which is a distillation column with a partial condenser and kettle reboiler. Vapors from the flash drum and the overhead vapors from the stripping column are sent to the Claus sulfur recovery unit to be treated with other sour gas streams. A portion of the bottoms product exiting the stripper is sent to the wastewater treatment plant and the remainder is recycled back to the syngas water scrubber.

Acid Gas Removal

H_2S and CO_2 are removed with a dual-stage Rectisol[®] unit. The Rectisol[®] AGR process was specified primarily because the FT synthesis catalyst requires a sulfur (H_2S and COS) level below 100 parts per billion by volume (ppbv) to maintain an adequate catalyst lifetime.

Sulfur Recovery Unit

The purpose of the SRU is to treat the acid gas from the Rectisol[®] unit and sour gas streams from the sour water strippers to make an effluent gas acceptable for venting to the atmosphere or burning.

For this study, the SRU is a Claus bypass type SRU utilizing oxygen instead of air. The Claus plant produces molten sulfur by converting approximately one third of the H₂S in the feed to SO₂, then reacting the H₂S and SO₂ to sulfur and water. The combined Claus technology and tail gas recycle to Rectisol[®] process results in an overall sulfur recovery exceeding 99%.

Fischer Tropsch Synthesis and Product Upgrade

The FT process and product upgrade block is shown in the Figure 4-2. Table 4-2 lists stream flows in this block, As stated in the previous sections, this FT synthesis process uses a cobalt-catalyzed, low-temperature FT process and two slurry-bed reactors in series to achieve about 85% conversion. The syngas is cooled after each reactor to condense out the hydrocarbons. The unconverted syngas leaving the FT block is sent for further product recovery. This low temperature product recovery step necessitates removal of CO₂ from the syngas stream to avoid operational issues due to CO₂ solidification in the lines. Hence, an activated amine based CO₂ removal system is required to remove CO₂ from this syngas. To recover hydrogen for the hydrotreating and hydrocracking reactions, syngas exiting the FT process, after the hydrocarbon recovery, is sent to a PSA system for hydrogen recovery. The tail gas from the PSA system is used as the fuel gas for power generation.

Figure 4-1
Case 5A Reference Plant – Overall BFD

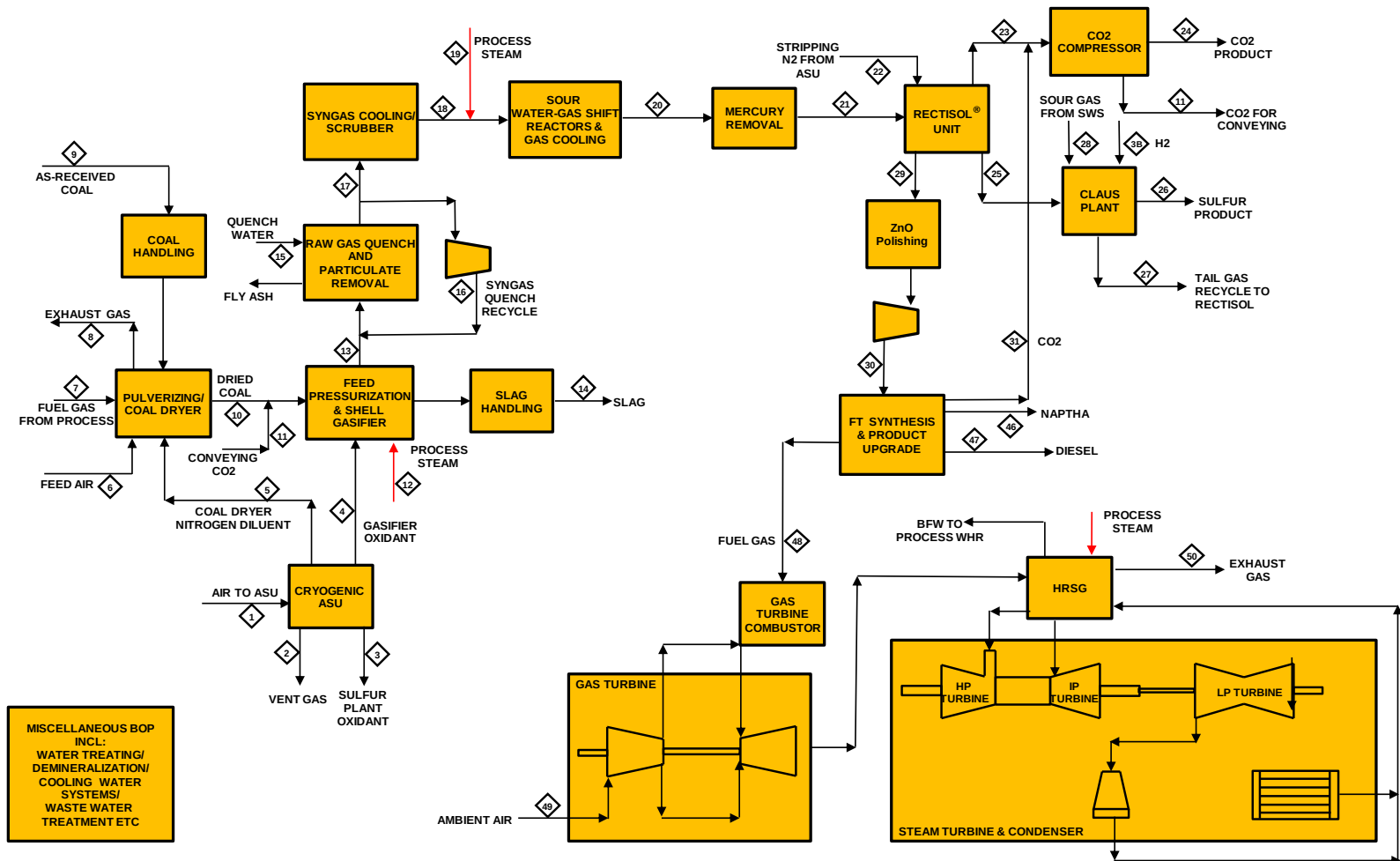


Table 4-1
Case 5A Reference Plant – Overall Stream Table

	1	2	3	3B	4	5	6	7	8	9
	Air to ASU	Vent Gas from ASU	Sulfur Plant Oxidant	Hydrogen to Sulfur Plant	Gasifier Oxidant	Coal Dryer Nitrogen Diluent	Air to Coal Dryer	Fuel Gas to Coal Dryer	Exhaust Gas from Coal Dryer	As-Received Coal
Mole Frac										
H2	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.0000	0.1582	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2290	0.0000	0.0000
CO2	0.0003	0.0077	0.0000	0.0000	0.0000	0.0000	0.0003	0.0073	0.0101	0.0000
CH4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1587	0.0000	0.0000
H2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N2	0.7732	0.7077	0.0178	0.0000	0.0178	0.9919	0.7732	0.1747	0.8963	0.0000
AR	0.0092	0.0225	0.0318	0.0000	0.0318	0.0023	0.0092	0.1496	0.0048	0.0000
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0099	0.0714	0.0000	0.0000	0.0000	0.0004	0.0099	0.0025	0.0775	0.0000
O2	0.2074	0.1906	0.9504	0.0000	0.9504	0.0054	0.2074	0.0000	0.0113	0.0000
SO2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Cl2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0027	0.0000	0.0000
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0212	0.0000	0.0000
C3H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0202	0.0000	0.0000
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0259	0.0000	0.0000
C4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0174	0.0000	0.0000
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0009	0.0000	0.0000
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0197	0.0000	0.0000
IC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0119	0.0000	0.0000
S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
GYPSUM	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
LIME	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mole Flow lbmol/hr		8,513	775	29	40,525	80,240	10,103	1,258	97,759	
Mass Flow lb/hr		242,213	24,940	59	1,304,140	2,251,410	291,540	32,808	2,688,530	1,750,520
Temperature F	59	89.6	90	85	241	70	59	94	180	59
Pressure psia	15	190	125	15	940	18	15	15	15	15
Solid Frac	0	0	0	0	0	0	0	0	0	1
Enthalpy Btu/hr	-2.44E+08	-8.54E+07	4.44E+04	-3.39E+03	4.09E+07	-7.59E+06	-1.22E+07	-2.62E+07	-8.84E+08	-1.58E+09
Average MW	28.856	28.451	32.181	2.016	32.181	28.058	28.856	26.087	27.502	

Table 4-1 (cont'd)
Case 5A Reference Plant – Overall Stream Table

	10	11	12	13	14	15	17	18	19	20
	Dried Coal	Conveying CO2	Gasifier Steam	Raw Syngas	Slag	Quench Water	Syngas to Scrubber	Syngas to Shift	Shift Steam	Shifted Syngas
Mole Frac										
H2	0.0000	0.0000	0.0000	0.2894	0.0000	0.0000	0.1831	0.2024	0.0000	0.4762
CO	0.0000	0.0000	0.0000	0.6093	0.0000	0.0000	0.3855	0.4261	0.0000	0.2323
CO2	0.0000	1.0000	0.0000	0.0263	0.0000	0.0000	0.0167	0.0184	0.0000	0.2690
CH4	0.0000	0.0000	0.0000	0.0003	0.0000	0.0000	0.0002	0.0002	0.0000	0.0002
H2S	0.0000	0.0000	0.0000	0.0082	0.0000	0.0000	0.0052	0.0057	0.0000	0.0067
COS	0.0000	0.0000	0.0000	0.0007	0.0000	0.0000	0.0004	0.0005	0.0000	0.0003
NH3	0.0000	0.0000	0.0000	0.0010	0.0000	0.0000	0.0006	0.0001	0.0000	0.0000
HCL	0.0000	0.0000	0.0000	0.0009	0.0000	0.0000	0.0006	0.0000	0.0000	0.0000
N2	0.0000	0.0000	0.0000	0.0093	0.0000	0.0000	0.0059	0.0065	0.0000	0.0074
AR	0.0000	0.0000	0.0000	0.0084	0.0000	0.0000	0.0053	0.0059	0.0000	0.0066
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0000	0.0000	1.0000	0.0460	0.0000	1.0000	0.3963	0.3343	1.0000	0.0012
O2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Cl2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
GYPSUM	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
LIME	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mole Flow lbmol/hr		4,970	9,376	153,005	-	88,803	241,808	218,743	50,175	194,020
Mass Flow lb/hr	1,637,750	218,718	168,908	3,154,190	169,744	1,599,810	4,754,000	4,335,680	903,925	3,890,000
Temperature F	60	80	650	2600	140	420	450	375	600	95
Pressure psia	840	750	740	615	615	1200	585	575	650	510
Solid Frac	1	0	0	0	1	0	0	0	0	0
Enthalpy Btu/hr	-8.08E+08	-8.47E+08	-9.38E+08	-2.85E+09	1988530	-1.04E+10	-1.45E+10	-1.23E+10	-5.04E+09	-1.10E+10
Average MW		44.010	18.015	20.615		18.015	19.660	19.821	18.015	20.049

Table 4-1 (cont'd)
Case 5A Reference Plant – Overall Stream Table

	21	22	23	25	26	27	28	29	30
	Syngas to Rectisol®	Stripping N2 from ASU	Total Captured CO2	Acid Gas from Rectisol®	Sulfur	Tail Gas to Rectisol®	Sour Gas from Stripper	Clean Syngas from Rectisol®	Treated Syngas to FT
Mole Frac									
H2	0.4762	0.0000	0.0034	0.0001	0.0000	0.0462	0.0483	0.6487	0.6487
CO	0.2323	0.0000	0.0082	0.0000	0.0000	0.0928	0.0970	0.3147	0.3147
CO2	0.2690	0.0000	0.9543	0.4160	0.0000	0.7090	0.7416	0.0166	0.0166
CH4	0.0002	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0003	0.0003
H2S	0.0067	0.0000	0.0000	0.5608	0.0001	0.0179	0.0187	0.0000	0.0000
COS	0.0003	0.0000	0.0000	0.0217	0.0000	0.0000	0.0000	0.0000	0.0000
NH3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N2	0.0074	1.0000	0.0176	0.0013	0.0000	0.0830	0.0869	0.0106	0.0106
AR	0.0066	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0091	0.0091
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0012	0.0000	0.0165	0.0000	0.0000	0.0511	0.0076	0.0000	0.0000
O2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Cl2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
S	0.0000	0.0000	0.0000	0.0000	0.9999	0.0000	0.0000	0.0000	0.0000
GYPSUM	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
LIME	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mole Flow lbmol/hr	194,020	882	52,206	2,369	1,373	1,369	1,309	142,253	142,253
Mass Flow lb/hr	3,890,000	24,704	2,246,160	91,820	44,032	51,700	50,617	1,638,430	1,638,430
Temperature F	95	100	56	60.61978	299.617975	120	95	88	200
Pressure psia	500	500	15	32	15	19	500	478	371
Solid Frac	0	0	0	0	0	37.75237	38.65898	11.51773	11.51773
Enthalpy Btu/hr	-1.10E+10	5.55E+04	-8.56E+09	-1.82E+08	1.00E+02	-1.78E+08	-1.55E+07	-2.52E+09	-2.41E+09
Average MW	20.049	28.013	43.025	38.766	0.000	37.972	21.090	11.518	11.518

Figure 4-2
Case 5A Reference Plant –FT Process BFD

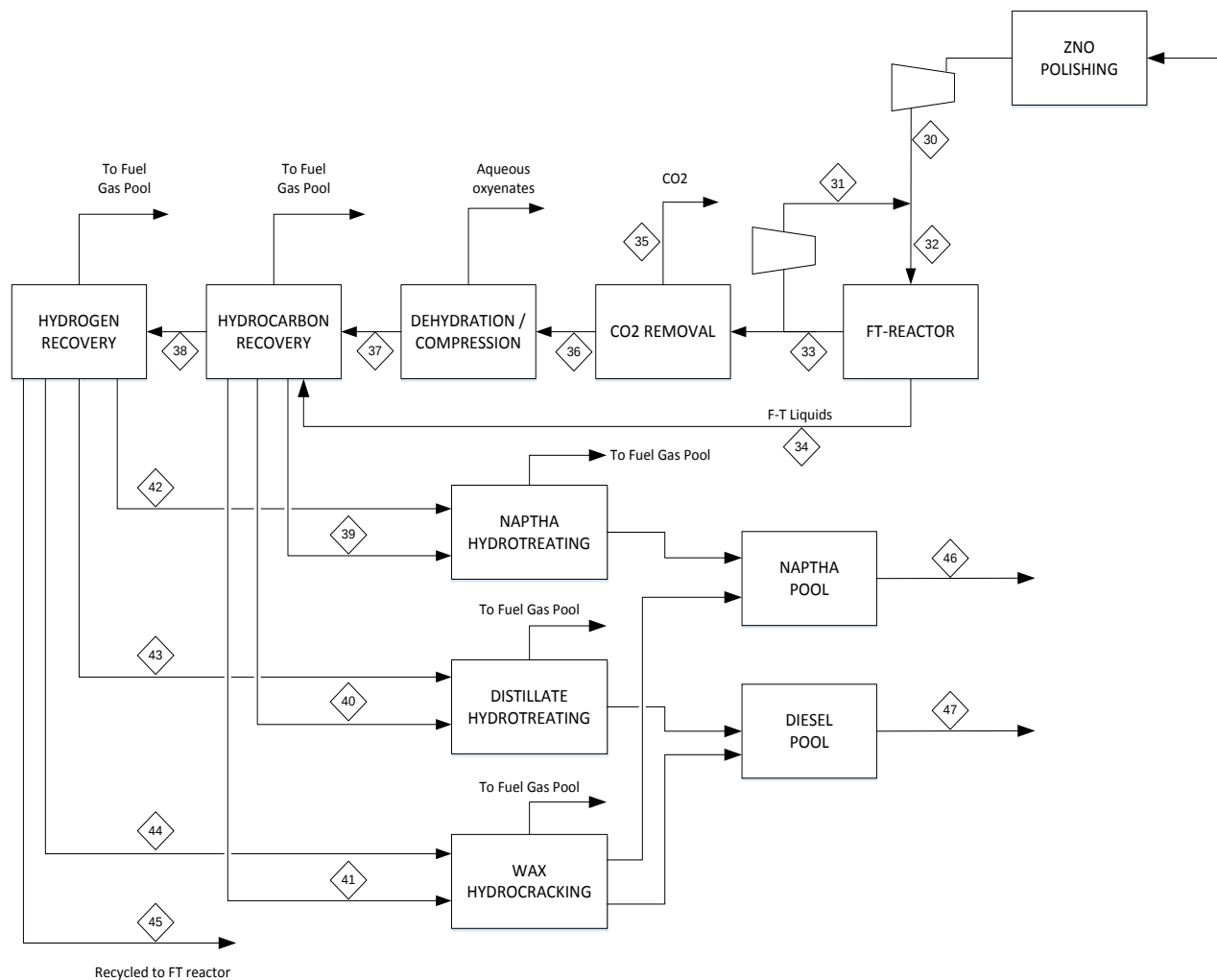


Table 4-2
Case 5A Reference Plant –FT Process Stream Table

	30	31	32	33	34	35	36	37	38
Component Mole Fraction									
O2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2	0.01057	0.09466	0.03060	0.09466	0.00229	0.00000	0.11060	0.11102	0.11686
CO	0.31474	0.12373	0.26298	0.12373	0.00501	0.00000	0.14457	0.14512	0.15238
CO2	0.01661	0.14487	0.04716	0.14487	0.02525	1.00000	0.00085	0.00085	0.00059
H2	0.64872	0.42438	0.60125	0.42438	0.00580	0.00000	0.49586	0.49775	0.52581
H2O	0.00000	0.00325	0.00078	0.00325	0.05503	0.00000	0.00380	0.00000	0.00000
NH3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCN	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H2S	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
AR	0.00906	0.08087	0.02616	0.08087	0.00300	0.00000	0.09449	0.09485	0.09885
CH4	0.00029	0.08298	0.02018	0.08298	0.00468	0.00000	0.09696	0.09733	0.09927
C2H4	0.00000	0.00144	0.00035	0.00144	0.00026	0.00000	0.00168	0.00169	0.00111
C2H6	0.00000	0.00747	0.00180	0.00747	0.00187	0.00000	0.00873	0.00877	0.00387
C3H6	0.00000	0.00992	0.00238	0.00992	0.00653	0.00000	0.01159	0.01163	0.00087
C3H8	0.00000	0.00450	0.00108	0.00450	0.00338	0.00000	0.00526	0.00528	0.00031
IC4H8	0.00000	0.00036	0.00009	0.00036	0.00069	0.00000	0.00042	0.00042	0.00000
C4H8	0.00000	0.00720	0.00173	0.00720	0.01396	0.00000	0.00841	0.00844	0.00005
IC4H10	0.00000	0.00017	0.00004	0.00017	0.00029	0.00000	0.00020	0.00020	0.00000
C4H10	0.00000	0.00318	0.00077	0.00318	0.00718	0.00000	0.00372	0.00373	0.00002
C5H10	0.00000	0.00222	0.00053	0.00222	0.01234	0.00000	0.00259	0.00260	0.00000
NC5H12	0.00000	0.00383	0.00092	0.00383	0.02563	0.00000	0.00448	0.00449	0.00000
IC5H12	0.00000	0.00049	0.00012	0.00049	0.00248	0.00000	0.00057	0.00057	0.00000
C6H12	0.00000	0.00105	0.00025	0.00105	0.01702	0.00000	0.00123	0.00123	0.00000
NC6H14	0.00000	0.00171	0.00041	0.00171	0.03398	0.00000	0.00200	0.00201	0.00000
IC6H14	0.00000	0.00024	0.00006	0.00024	0.00349	0.00000	0.00028	0.00028	0.00000
C7H14	0.00000	0.00037	0.00009	0.00037	0.01893	0.00000	0.00044	0.00044	0.00000
C7H16	0.00000	0.00067	0.00016	0.00067	0.04057	0.00000	0.00079	0.00079	0.00000
C8H16	0.00000	0.00011	0.00003	0.00011	0.01852	0.00000	0.00013	0.00013	0.00000
C8H18	0.00000	0.00020	0.00005	0.00020	0.03924	0.00000	0.00023	0.00023	0.00000
C9H18	0.00000	0.00003	0.00001	0.00003	0.01721	0.00000	0.00004	0.00004	0.00000
C9H20	0.00000	0.00005	0.00001	0.00005	0.03631	0.00000	0.00006	0.00006	0.00000
C10H20	0.00000	0.00001	0.00000	0.00001	0.01571	0.00000	0.00001	0.00001	0.00000
C10H22	0.00000	0.00001	0.00000	0.00001	0.03310	0.00000	0.00002	0.00002	0.00000
C11H22	0.00000	0.00000	0.00000	0.00000	0.00146	0.00000	0.00000	0.00000	0.00000

Table 4-2 (cont'd)
Case 5A Reference Plant – FT Process Stream Table

	30	31	32	33	34	35	36	37	38
Component Mole Fraction									
C11H24	0.00000	0.00001	0.00000	0.00001	0.04284	0.00000	0.00001	0.00001	0.00000
C12H24	0.00000	0.00000	0.00000	0.00000	0.00133	0.00000	0.00000	0.00000	0.00000
C12H26	0.00000	0.00000	0.00000	0.00000	0.03883	0.00000	0.00000	0.00000	0.00000
C13H26	0.00000	0.00000	0.00000	0.00000	0.00120	0.00000	0.00000	0.00000	0.00000
C13H28	0.00000	0.00000	0.00000	0.00000	0.03520	0.00000	0.00000	0.00000	0.00000
C14H28	0.00000	0.00000	0.00000	0.00000	0.00109	0.00000	0.00000	0.00000	0.00000
C14H30	0.00000	0.00000	0.00000	0.00000	0.03188	0.00000	0.00000	0.00000	0.00000
C15H30	0.00000	0.00000	0.00000	0.00000	0.00099	0.00000	0.00000	0.00000	0.00000
C15H32	0.00000	0.00000	0.00000	0.00000	0.02888	0.00000	0.00000	0.00000	0.00000
C16H32	0.00000	0.00000	0.00000	0.00000	0.00089	0.00000	0.00000	0.00000	0.00000
C16H34	0.00000	0.00000	0.00000	0.00000	0.02616	0.00000	0.00000	0.00000	0.00000
C17H34	0.00000	0.00000	0.00000	0.00000	0.00081	0.00000	0.00000	0.00000	0.00000
C17H36	0.00000	0.00000	0.00000	0.00000	0.02370	0.00000	0.00000	0.00000	0.00000
C18H36	0.00000	0.00000	0.00000	0.00000	0.00073	0.00000	0.00000	0.00000	0.00000
C18H38	0.00000	0.00000	0.00000	0.00000	0.02146	0.00000	0.00000	0.00000	0.00000
C19H38	0.00000	0.00000	0.00000	0.00000	0.00066	0.00000	0.00000	0.00000	0.00000
C19H40	0.00000	0.00000	0.00000	0.00000	0.01944	0.00000	0.00000	0.00000	0.00000
C20H40	0.00000	0.00000	0.00000	0.00000	0.00060	0.00000	0.00000	0.00000	0.00000
C20H42	0.00000	0.00000	0.00000	0.00000	0.01761	0.00000	0.00000	0.00000	0.00000
C21OP	0.00000	0.00000	0.00000	0.00000	0.01688	0.00000	0.00000	0.00000	0.00000
C22OP	0.00000	0.00000	0.00000	0.00000	0.01565	0.00000	0.00000	0.00000	0.00000
C23OP	0.00000	0.00000	0.00000	0.00000	0.01450	0.00000	0.00000	0.00000	0.00000
C24OP	0.00000	0.00000	0.00000	0.00000	0.01344	0.00000	0.00000	0.00000	0.00000
C25OP	0.00000	0.00000	0.00000	0.00000	0.01246	0.00000	0.00000	0.00000	0.00000
C26OP	0.00000	0.00000	0.00000	0.00000	0.01155	0.00000	0.00000	0.00000	0.00000
C27OP	0.00000	0.00000	0.00000	0.00000	0.01070	0.00000	0.00000	0.00000	0.00000
C28OP	0.00000	0.00000	0.00000	0.00000	0.00992	0.00000	0.00000	0.00000	0.00000
C29OP	0.00000	0.00000	0.00000	0.00000	0.00919	0.00000	0.00000	0.00000	0.00000
C30WAX	0.00000	0.00000	0.00000	0.00000	0.11958	0.00000	0.00000	0.00000	0.00000
OXVAP	0.00000	0.00000	0.00000	0.00000	0.00224	0.00000	0.00000	0.00000	0.00000
OXHC	0.00000	0.00002	0.00000	0.00002	0.01838	0.00000	0.00002	0.00002	0.00000
OXH2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
GASOLINE	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
DIESEL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Mole Flow (LBMOL/HR)	142,253	46,174	191,974	61,978	2,590	2,278	13,526	13,475	12,747
Mass Flow (LB/HR)	1,638,430	939,568	2,585,150	1,261,170	555,176	100,263	221,334	220,409	183,194
Temperature (F)	200	149	184	100	285	100	100	100	85
Pressure (PSIA)	371	378	371	284	75	275	275	430	395

Table 4-2 (cont'd)
Case 5A Reference Plant – FT Process Stream Table

	39	40	41	42	43	44	45	46	47
Component Mole Fraction									
O2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CO2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H2	0.00000	0.00000	0.00000	1.00000	1.00000	1.00000	1.00000	0.00000	0.00000
H2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NH3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCN	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H2S	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
AR	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C2H4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C2H6	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H6	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H8	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
IC4H8	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H8	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
IC4H10	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H10	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H10	0.06621	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NC5H12	0.12550	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.07034	0.00000
IC5H12	0.01394	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.03068	0.00000
C6H12	0.06001	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NC6H14	0.11374	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.15204	0.00000
IC6H14	0.01263	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.05372	0.00000
C7H14	0.05433	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H16	0.11441	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.43992	0.00000
C8H16	0.04918	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H18	0.10356	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.06050	0.00000
C9H18	0.04455	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H20	0.09381	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H20	0.04037	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H22	0.08501	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C11H22	0.00000	0.00509	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000

Table 4-2 (cont'd)
Case 5A Reference Plant – FT Process Stream Table

	39	40	41	42	43	44	45	46	47
Component Mole Fraction									
C11H24	0.00000	0.14929	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C12H24	0.00000	0.00461	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C12H26	0.00000	0.13523	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C13H26	0.00000	0.00418	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C13H28	0.00000	0.12255	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C14H28	0.00000	0.00379	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C14H30	0.00000	0.11100	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C15H30	0.00000	0.00343	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C15H32	0.00000	0.10056	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C16H32	0.00000	0.00311	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C16H34	0.00000	0.09108	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C17H34	0.00000	0.00282	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C17H36	0.00000	0.08251	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C18H36	0.00000	0.00255	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C18H38	0.00000	0.07472	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C19H38	0.00000	0.00231	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C19H40	0.00000	0.06768	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H40	0.00000	0.00000	0.00238	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H42	0.00000	0.00000	0.06986	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C21OP	0.00000	0.00000	0.06697	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C22OP	0.00000	0.00000	0.06208	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C23OP	0.00000	0.00000	0.05753	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C24OP	0.00000	0.00000	0.05333	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C25OP	0.00000	0.00000	0.04942	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C26OP	0.00000	0.00000	0.04580	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C27OP	0.00000	0.00000	0.04245	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C28OP	0.00000	0.00000	0.03934	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C29OP	0.00000	0.00000	0.03646	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C30WAX	0.00000	0.00000	0.47438	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
OXVAP	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
OXHC	0.02274	0.03348	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
OXH2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
GASOLINE	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.19280	0.00000
DIESEL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.00000
Mole Flow (LBMOL/HR)	1.011	744	653	502	40	1.273	3.547	1.491	1.832
Mass Flow (LB/HR)	103.792	148.774	305.189	1.012	81	2.565	7.151	150.157	389.179
Temperature (F)	255	255	255	100	100	602	85	100	100
Pressure (PSIA)	50	50	50	50	50	50	395	50	50

4.2 SPARING PHILOSOPHY

The sparing philosophy for the key process blocks for the Case 5A CTL plant is provided below. The design has:

- Eight trains of gasification, including dry coal feed system, and Shell gasifier
- Four Rectisol[®] trains (4 x 25%)
- Two Claus trains (2 x 50%)

4.3 PERFORMANCE RESULTS

The RTI-modeled Case 5A reference plant with CO₂ capture consumes 21,006 tpd of Illinois #6 coal at the Midwestern site and produces about 50,000 bbl/day of FT liquid product, containing 70% FT diesel and 30% FT Naphtha. Overall performance for the Case 5A reference plant is summarized in Table 4-3, which includes auxiliary power requirements.

Table 4-3
Case 5A Reference Plant – Performance Summary Table

Power Cycle Power Generation	Electrical Load, kWe
Gas Turbine Power	122,547
Steam Turbine Power	403,151
Syngas Turboexpansion Power	7,326
Total	533,024
CTL Auxiliary Load Description	Electrical Load, kWe
Coal Handling	1,760
Coal Milling	8,160
Slag Handling	2,080
ASU Auxiliaries	3,810.80
ASU Main Air Compressor	224,342
Oxygen Compressor	36,324
Quench Water Pump	2,299
Syngas Recycle Compressor	2,920
Scrubber Pumps	1,360
Rectisol® Auxiliary	49,120
Claus Plant Auxiliary	1,000
Claus Plant TG Compressor	840
CO ₂ Compressor Auxiliary	78,410
FT Power Requirement	30,390
Miscellaneous BOP	13,959
Circulating Water Pumps	14,640
Ground Water Pumps	800
Cooling Tower Fans	7,665
Gas Turbine Auxiliaries	4,000
Steam Turbine Auxiliaries	400
Transformer Losses	3,144
Condensate Pumps	200
Boiler Feed Water Pumps	4,065
Total	491,689
NET PLANT POWER, kWe	41,334

Table 4-4 shows the carbon balance for the Case 5A reference plant. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal feedstock. Carbon in the air is not part of the carbon capture equation, but is not neglected in the balance since the model accounts for the air components throughout. Carbon leaves the CFTL process plant as unburned

carbon in the slag, CO₂ in the dryer exhaust gas, ASU vent gas, FT liquids product, carbon in the sour water, the Rectisol[®] AGR CO₂ capture product and the post FT AACRP CO₂ capture product. Carbon in the F-T liquids is considered as product, not emissions. The carbon capture efficiency is defined as the amount of carbon in the product streams, which include the F-T liquids and the dried and compressed CO₂ products from the Rectisol[®] AGR and AACRP CO₂ (post FT) capture processes, relative to the amount of carbon in the coal feedstock less carbon contained in the slag. For Case 5A, the carbon capture efficiency is 92.5%.

Table 4-4
Case 5A Reference Plant – Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	1,115,957	
ASU Air	789	
Coal Dryer Air	36	
Carbon in Sour water		1,439
Carbon in LTGC condensate		-
ASU Vent		789
Exhaust Gas to Stack		81,751
CO ₂ (post FT)		27,361
Carbon in Slag		5,579
Sulfur Product		-
FT Naphtha		125,994
FT Diesel		330,058
Rectisol [®] CO ₂ Product		543,812
Total	1,116,782	1,116,782

Table 4-5 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur emitted in the coal dryer exhaust gas, and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible. There is no net sulfur emission from the Case 5A reference plant.

Table 4-5
Case 5A Reference Plant – Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	43,938	
Coal Dryer Exhaust Gas		-
Exhaust Gas to Stack		-
CO ₂ (post FT)		-
Sulfur Product		43,924
FT Naphtha		-
FT Diesel		-

Rectisol® CO ₂ Product		14
Total	43,938	43,938

Table 4-6 shows the overall water balance for the Case 5A reference plant. Raw water withdrawal is the amount of raw water consumed by the plant. Some water is internally recycled and consumed by the slag handling process, Quench and syngas scrubber and as makeup to the cooling tower.

Table 4-6 Case 5A Reference Plant – Overall Water Balance

Water Usage by Area	Water Demand		Internal Recycle		Raw water Withdrawal		Process Water Discharge		Raw Water Consumption	
	m ³ /min	gpm	m ³ /min	gpm	m ³ /min	gpm	m ³ /min	gpm	m ³ /min	gpm
Slag Handling	1.7	456	1.7	456	-	-	-	-	-	-
Quench/Wash	12.1	3,193	10.4	2,738	1.7	456	-	-	1.7	456
AGR Makeup Water	-	-	-	-	-	-	-	-	-	-
SWS Blowdown	-	-	-	-	-	-	0.1	30	(0.1)	(30)
FT-Area	0.2	42	-	-	0.2	42	-	-	0.2	42
Condenser Makeup	8.6	2,280	-	-	8.6	2,280	-	-	8.6	2,280
Gasifier Steam	1.3	337	-	-	1.3	337	-	-	1.3	337
Shift Steam	6.8	1,804	-	-	6.8	1,804	-	-	6.8	1,804
BFW Makeup	0.5	138	-	-	0.5	138	-	-	0.5	138
Cooling Tower	57.7	15,247	-	-	57.7	15,247	14.4	3,805	43.3	11,442
BFW Blowdown	-	-	0.5	138	(0.5)	(138)	-	-	(0.5)	(138)
SWS Blowdown	-	-	1.1	302	(1.1)	(302)	-	-	(1.1)	(302)
TOTAL	80.4	21,217		3,633	66.6	17,583		3,835	52.1	13,749

4.4 CAPITAL COST

Table 4-7 shows the cost breakdown of the Case 5A reference plant, consistent with the code of accounts format as expressed in the CTL Study. Syngas cleanup technology blocks were costed in detail, whereas the other accounts were scaled from various DOE studies following DOE scaling guidelines.

The key accounts are shown with more detail to include the various subaccounts and provide more clarity to the major cost differences among the cases.

Table 4-7
Case 5A Reference Plant – Total Plant Cost Summary

Case 5a: Shell with Rectisol® AGR CFTL Plant				Coal Feed, lb/hr 1,750,518 Coal HHV, Btu/lb 13,126 (dry basis)				Plant Size Cost Base		50,000 barrels per day of FT-liquids Jun 2011 (\$x1000)		
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost \$	Eng'g CM H.O & Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect				Process	Project	\$	
1	COAL & SORBENT HANDLING	\$38,319	\$6,719	\$29,308	\$0	\$0	\$74,346	\$6,589	\$0	\$16,187	\$97,122	
2	COAL & SORBENT PREP & FEED											
2.3	Dry Coal Injection System	\$189,497	\$2,184	\$17,357	\$0	\$0	\$209,038	\$17,410	\$0	\$45,290	\$271,738	
2.x	Other Coal & Sorbent Prep & Feed Systems	\$130,490	\$22,642	\$35,141	\$0	\$0	\$188,273	\$15,985	\$0	\$40,851	\$245,109	
	SUBTOTAL 2.	\$319,987	\$24,826	\$52,498	\$0	\$0	\$397,311	\$33,395	\$0	\$86,141	\$516,847	
3	FEEDWATER & MISC BOP SYSTEMS	\$15,602	\$9,762	\$17,268	\$0	\$0	\$42,632	\$3,950	\$0	\$11,365	\$57,947	
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries	\$699,517	\$0	\$301,659	\$0	\$0	\$1,001,176	\$88,113	\$137,204	\$188,698	\$1,415,191	
4.3	ASU/Oxidant Compression	\$532,888	\$0 w/equip.		\$0	\$0	\$532,888	\$49,793	\$0	\$58,268	\$640,950	
4.4	LT Heat Recovery & FG Saturation	\$90,428	\$0	\$28,713	\$0	\$0	\$119,140	\$11,257	\$0	\$26,079	\$156,475	
4.6	Flare Stack System	\$0	\$3,441	\$1,390	\$0	\$0	\$4,832	\$451	\$0	\$1,056	\$6,340	
4.9	Gasification Foundations	\$0	\$41,667	\$24,849	\$0	\$0	\$66,516	\$6,131	\$0	\$18,162	\$90,810	
	SUBTOTAL 4.	\$1,322,833	\$45,108	\$356,611	\$0	\$0	\$1,724,553	\$155,745	\$137,204	\$292,263	\$2,309,765	
5A	GAS CLEANUP & PIPING											
5A.1a	Rectisol® System	\$283,240	\$0	\$230,161	\$0	\$0	\$513,401	\$48,442	\$102,680	\$132,905	\$797,427	
5A.1b	CO2 Removal System	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
5A.2	Elemental Sulfur Plant	\$24,556	\$4,786	\$31,465	\$0	\$0	\$60,807	\$5,775	\$0	\$13,317	\$79,899	
5A.3	Mercury Removal	\$2,985	\$0	\$2,256	\$0	\$0	\$5,241	\$494	\$262	\$1,199	\$7,196	
5A.4	Shift Reactors	\$4,825	\$0	\$1,930	\$0	\$0	\$6,755	\$631	\$0	\$1,477	\$8,863	
5A.5	Particulate Removal	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
5A.6	Blowback Gas Systems	\$3,633	\$611	\$342	\$0	\$0	\$4,586	\$423	\$0	\$1,002	\$6,011	
5A.7	Fuel Gas Piping	\$0	\$1,733	\$1,134	\$0	\$0	\$2,867	\$250	\$0	\$624	\$3,741	
5A.9	HGCU Foundations	\$0	\$1,965	\$1,324	\$0	\$0	\$3,289	\$304	\$0	\$1,078	\$4,671	
5A.10	Syngas Conditioning	\$26,360	\$18,425	\$21,342	\$0	\$0	\$66,127	\$6,282	\$0	\$14,482	\$86,891	
	SUBTOTAL 5.	\$345,599	\$27,520	\$289,954	\$0	\$0	\$663,073	\$62,601	\$102,942	\$166,083	\$994,699	
5AA	F-T SYNTHESIS AND PRODUCT UPGRADE											
5AA.1	FT Synthesis*	\$148,753 w/equip.	w/equip.	\$0	\$0	\$0	\$148,753	\$14,280	\$40,163	\$50,800	\$253,996	
5AA.2	Amine CO2 Absorption	\$7,937 w/equip.	w/equip.	\$0	\$0	\$0	\$7,937	\$749	\$1,587	\$2,055	\$12,328	
5AA.4	Overhead Recovery	\$8,345 w/equip.	w/equip.	\$0	\$0	\$0	\$8,345	\$801	\$2,253	\$2,850	\$14,249	
5AA.5	Product Separation	\$36,466 w/equip.	w/equip.	\$0	\$0	\$0	\$36,466	\$3,501	\$9,846	\$12,453	\$62,266	
5AA.6	Hydrogen Recovery	\$12,370 w/equip.	w/equip.	\$0	\$0	\$0	\$12,370	\$1,188	\$3,340	\$4,224	\$21,122	
5AA.8	Naphtha Hydrotreater	\$13,250 w/equip.	w/equip.	\$0	\$0	\$0	\$13,250	\$1,272	\$3,578	\$4,525	\$22,625	
5AA.9	Diesel Hydrotreater	\$33,271 w/equip.	w/equip.	\$0	\$0	\$0	\$33,271	\$3,195	\$8,983	\$11,363	\$56,812	
5AA.10	Wax Hydrocracker	\$81,815 w/equip.	w/equip.	\$0	\$0	\$0	\$81,815	\$7,855	\$22,091	\$27,940	\$139,700	
5AA.11	Raw Fuel Gas Compressor, HP	\$24,034 w/equip.	w/equip.	\$0	\$0	\$0	\$24,034	\$2,242	\$0	\$5,255	\$31,531	
5AA.12	Fuel Gas to GT Compressor, HP	\$6,334 w/equip.	w/equip.	\$0	\$0	\$0	\$6,334	\$591	\$0	\$1,385	\$8,310	
	SUBTOTAL 5AA.	\$372,575	\$0	\$0	\$0	\$0	\$372,575	\$35,672	\$91,841	\$122,850	\$622,938	
5B	CO2 Compression & Drying	\$85,515	\$0	\$32,238	\$0	\$0	\$117,753	\$10,980	\$0	\$25,748	\$154,481	
	SUBTOTAL 5B.	\$85,515	\$0	\$32,238	\$0	\$0	\$117,753	\$10,980	\$0	\$25,748	\$154,481	
6	COMBUSTION TURBINE/ACCESSORIES	\$48,358	\$394	\$3,814	\$0	\$0	\$52,566	\$11,379	\$12,131	\$15,129	\$91,205	
7	HRSG, DUCTING & STACK	\$19,039	\$1,670	\$5,421	\$0	\$0	\$26,130	\$2,409	\$0	\$3,324	\$31,863	
8	STEAM TURBINE GENERATOR	\$53,595	\$1,455	\$19,016	\$0	\$0	\$74,067	\$6,451	\$0	\$10,752	\$91,270	
9	COOLING WATER SYSTEM	\$16,095	\$15,458	\$13,779	\$0	\$0	\$45,332	\$4,067	\$0	\$10,474	\$59,874	
10	ASH/SPENT SORBENT HANDLING SYS	\$51,892	\$3,622	\$25,536	\$0	\$0	\$81,050	\$7,576	\$0	\$9,611	\$98,237	
11	ACCESSORY ELECTRIC PLANT	\$40,827	\$23,823	\$41,547	\$0	\$0	\$106,197	\$9,001	\$0	\$22,888	\$138,087	
12	INSTRUMENTATION & CONTROL	\$15,839	\$3,163	\$10,261	\$0	\$0	\$29,262	\$2,604	\$1,463	\$5,588	\$38,918	
13	IMPROVEMENTS TO SITE	\$8,281	\$5,915	\$25,247	\$0	\$0	\$39,443	\$3,905	\$0	\$13,004	\$56,352	
14	BUILDINGS & STRUCTURES	\$0	\$17,750	\$18,413	\$0	\$0	\$36,164	\$3,214	\$0	\$6,288	\$45,666	
	CALCULATED TOTAL COST	\$2,754,358	\$187,186	\$940,912	\$0	\$0	\$3,882,456	\$359,538	\$345,580	\$817,695	\$5,405,269	

Table 4-8 shows the calculation and addition of owner's costs to determine the total overnight cost (TOC), which is used to calculate the F-T products RSP.

The estimated TOC of the Case 5A reference plant in 2011 dollars is \$6,771 MM.

Table 4-8
Case 5A Reference Plant – Total Overnight Cost Breakdown

Owner's Costs	\$ x \$1,000
Preproduction Costs	
6 months All Labor	\$48,827
1 Month Maintenance Materials	\$8,458
1 Month Non-Fuel Consumables	\$4,596
1 Month Waste Disposal	\$1,617
25% of 1 Months Fuel Cost at 100% CF	\$10,958
2% of TPC	\$108,105
Total	\$182,561
Inventory Capital	
60-day supply of fuel at 100% CF	\$86,461
60-day supply of non-fuel consumables at 100% CF	\$9,065
0.5% of TPC (spare parts)	\$27,026
Total	\$122,552
Initial Cost for Catalyst and Chemicals	\$121,848
Land	\$900
Other Owner's Cost	\$774,948
Financing Costs	\$139,491
Total Owner's Costs	\$1,337,825
Total Overnight Costs (TOC)	\$6,771,090
TASC Multiplier	1.147
Total As-Spent Cost (TASC)	\$7,766,440

4.5 OPERATING COSTS

Table 4-9 shows the operating cost breakdown for the Case 5A reference plant.

Table 4-9
Case 5A Reference Plant – Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Shell CFTL w/Rectisol®-based AGR and CO2 Capture				
Plant Size (bbl/day FT Liquids)	50,000				
Primary/Secondary Fuel:	Illinois #6 Coal				
Design/Construction	5 years	Book Life (yrs):		20	
TPC (Plant Cost) Year	June 2011	TPI Year:		2016	
Capacity Factor (%)	90	CO2 Captured (TPD)		25,116	
OPERATING & MAINTENANCE LABOR					
Operating Labor					
Operating Labor Rate (base):	\$39.70 \$/hr				
Operating Labor Burden:	30.00 % of base				
Labor Overhead Charge	25.00 % of labor				
Operating Labor Requirements per Shift	units/mod	Total Plant			
Skilled Operator	2.0	8.0			
Operator	10.0	40.0			
Foreman	1.0	4.0			
Lab Tech's etc	3.0	12.0			
TOTAL Operating Jobs	16.0	64.0			
Annual Cost					
\$					
Annual Operating Labor Cost	\$28,934,630				
Maintenance Labor Cost	\$49,187,950				
Administration & Support Labor	\$19,530,645				
Property Taxes and Insurance	\$108,105,384				
TOTAL FIXED OPERATING COSTS	\$205,758,609				
VARIABLE OPERATING COSTS					
Maintenance Material Cost					
					\$91,349,050
Consumables	Consumption	Unit	Initial Fill		
	Initial	/Day	Cost	Cost	
Water(1000 gallons)	0	12,660	1.67	\$0	\$6,961,249
Chemicals					
MU & WT Chem (lb)	0	75417	0.27	\$0	\$6,635,740
Carbon (Hg Removal) (lb)	301,013	515	1.63	\$489,217	\$275,185
FT Catalyst (lbs)	4,346,525	4347	20.00	\$86,930,497	\$28,562,190
Water Gas Shift Catalyst (ft3)	11,498	6	771.99	\$8,876,622	\$1,597,792
Methanol (tons)	3,020	4	300.00	\$906,000	\$413,910
Amine Solution (gal)	13,785	4	23.38	\$322,298	\$29,007
Hydrotreating catalyst (ft3)	773	0.71	700.00	\$540,835	\$162,187
Zinc Oxide (ton)	689	1.89	7272.73	\$5,009,455	\$4,508,509
Claus Catalyst (ft3)	w/equip	7	203.15	\$0	\$486,163
Subtotal Chemicals				\$103,074,923	\$42,670,683
Other					
Butane (tons)	0	0.0	651.34	\$0	\$0
Gases, N2 etc./100scf	0	0	0.00	\$0	\$0
LP Steam (/1000 lbs)	0	0	0.00	\$0	\$0
Subtotal Other				\$0	\$0
Waste Disposal:					
Spent Mercury Catalyst (lb)	0	515	0.65	\$0	\$110,074
Flyash (ton)	0	0	0.00	\$0	\$0
Slag (ton)	0	2104	25.11	\$0	\$17,354,113
Subtotal Waste Disposal				\$0	\$17,464,187
By-products & Emissions					
Sulfur (tons)	0	0.0	0.00	\$0	\$0
Supplemental Electricity (MWh generated)	0	992	-60.00	\$0	-\$19,552,856
Subtotal By-Products				\$0	-\$19,552,856
TOTAL VARIABLE OPERATING COSTS				\$103,074,923	\$138,892,312
Coal (tons)	0	21,006	68.60	\$0	\$473,372,311

4.6 FISCHER-TROPSCH PRODUCT REQUIRED SELLING PRICE

Table 4-10 shows a summary of the power output, CAPEX, OPEX, and FT RSP for the Case 5A Plant. The Case 5A Plant FT Diesel COP is estimated to be \$134.3/bbl under the loan guarantee finance structure and \$156.0/bbl under the commercial fuels finance structure.

Table 4-10
Case 5A CTL Plant – Overall Performance and Economic Summary

Case	Case 5A	
Gasifier	Shell	
Sulfur and CO ₂ Removal	Dual Stage Rectisol®	
CAPEX, \$MM		
Total Installed Cost (TIC)	\$3,882	
Total Plant Cost (TPC)	\$5,405	
Total Overnight Cost (TOC)	\$6,771	
OPEX, \$MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC _{fix})	\$205.8	
Variable Operating Cost, less Fuel (OC _{var})	\$176.1	
Coal Fuel (OC _{fuel})	\$526.0	
Import/(Export) Power (OC _{power})	(\$21.7)	
Total OPEX	\$886.1	
Plant Output		
Crude FT Diesel Product, bbl/day	34,992	
Crude FT Naphtha Product, bbl/day	15,008	
Net Power Output, MWe	41.3	
Power Generation		
Gas Turbine Power	122.5	
Steam Turbine Power	403.2	
Syngas Turboexpansion Power	7.3	
Total Gross Power	533.0	
Net Power Generation/(Consumption), MWe	41.3	
Common Factors Used for RSP Calculation		
Capital Charge Factor (CCF)	0.1700	0.218
Capacity Factor (CF)	0.9	
Required Selling Price^{A,B,C}		
RSP FT Diesel, \$/bbl _{FTD}	134.3	156.0
RSP FT Naphtha, \$/bbl _{FTN}	94.2	109.2
Equivalent Crude Oil Price, \$/bbl _{ECO}	116.5	135.4
Equivalent Petroleum Diesel Price, \$/bbl _{EPD}	145.7	169.3
FT Diesel Required Selling Price Breakdown		
CAPEX	77.02	98.76
Fixed O&M	13.77	13.77
Variable O&M	10.60	10.60
Fuel	31.67	31.67
Power Generation	-1.31	-1.31
CO ₂ T&S	2.50	2.50
Total	134.25	156.00
CO₂ Capture		
CO ₂ Capture Product, tpd	25,116	
Total Carbon Captured (as % Carbon in Fuel)	92.5%	
Costs of CO₂ Captured^{A,B,C} (\$/ton)	13.92	15.00

^A Capacity factor assumed to be 90 percent

^B Values shown are for two financial structures

The first (lower value) is based on the loan guarantee finance structure

The second (higher value) is based on the commercial fuels finance structure

^C Includes CO₂ T&S

Case 5B has the plant configuration with Shell Gasifier and RTI WDP in the CTL plant. Case 5B has the same configuration as Case 5A, with the exception that the AGR Rectisol[®] process is replaced by RTI's WDP technology for sulfur removal and activated amine process for CO₂ removal (AACRP). The Claus process, in case 5A, for sulfur recovery is replaced by the DSRP in Case 5B. Syngas conversion to FT liquids follows the same process design for Case 5B, as Case 5A. Exhaust gas from the FT process is treated as the fuel gas for the gas turbine to generate power. Gas turbine is coupled with HRSG to generate power in the steam turbine. The combined power generated in the gas turbine and steam turbine is sufficient to meet the plant's auxiliary demands such that no power needs to be imported, and any excess power is exported for extra revenue.

The Case 5B plant is designed to capture CO₂ with a carbon capture efficiency of more than 90% of the carbon in the coal. The Case 5B plant operates with an annual on-stream capacity factor of 90%.

The Case 5B plant has the following characteristics that differentiate it from the Case 5A reference plant:

- Syngas leaving the gasifier that has been quenched and filtered enters the HCl guard bed, containing sodium bicarbonate, for chlorine removal (note that such a guard bed is only required for WDP for this case because of the high level of HCl found in syngas produced from Illinois #6 coal and would not normally be needed for most other coals).
- After the chlorine removal, syngas enters the RTI's WDP for sulfur removal at about 750°F instead of going through a low temperature scrubber as in Case 5A.
- After sulfur removal in the WDP, a portion of the treated syngas then enters the advanced fixed-bed water gas shift (AFWGS) process. Instead of the sour shift WGS process used in Case 5A, the Case 5B AFWGS process utilizes sweet shift catalyst which can operate at a higher temperature than the sour shift catalyst and at about half the cost.
- After the syngas is shifted to a 2:1 H₂/CO ratio and cooled, it enters the AACRP for CO₂ capture. Unlike Rectisol[®], which is a physical solvent, activated methyldiethanolamine is a chemical solvent. As ~99.9% of the sulfur compounds have been removed by the WDP process upstream already, the amine process only has to remove CO₂ and is less complicated than the two-stage Rectisol[®] setup. However, the AACRP unit does also capture ~99% of any residual sulfur left in the syngas following WDP along with the CO₂, resulting in an overall system total sulfur reduction of >99.99% (sub-ppmv total sulfur in the final cleaned syngas).
- In Case 5B, the DSRP replaces the Claus process, in Case 5A, for sulfur recovery. In the DSRP, sulfur leaving the WDP process in the form of SO₂ is reduced by a slipstream of shifted syngas, forming elemental sulfur, H₂S and COS. The elemental sulfur is

condensed while the remaining H₂S and COS are re-oxidized in the presence of air to form SO₂. The SO₂ is then removed in a lime scrubber, forming gypsum (CaSO₄·2H₂O). This combined approach for sulfur capture results in very low net SO₂ emissions.

The coal feed rate for Case 5B is expected to be the same as Case 5A, since Case 5B also utilizes the Shell gasifier operating under the same conditions. It therefore has the same cold gas efficiency and generates the same amount of raw gasifier syngas that is eventually converted to FT liquids product.

5.1 CASE 5B PROCESS DESCRIPTION

The overall BFD for the Case 5B plant is shown in Figure 5-2, with the accompanying stream flows shown in Table 5-1. The detailed BFD for the syngas conversion to F-T liquids block is shown in Figure 5-3 and the corresponding stream flows are provided in the Table 5.2. Additional descriptions of the CTL plant's various processes are provided below.

Coal Feed System

Same as Case 5A.

Shell Gasifier

Same as Case 5A.

Dry Solids Removal

Contrary to Case 5A, the raw syngas exiting the ceramic particulate filter enters an HCl guard bed for the removal of chlorides. The HCl guard bed uses sodium bicarbonate, which reacts with HCl forming NaCl, carbon dioxide and H₂O, for chloride removal. Chloride removal is achieved in four packed beds, which removes essentially all traces of entrained particles and chlorine. Note that such a guard bed is only required for WDP for this case because of the high level of HCl found in syngas produced from Illinois #6 coal and would not normally be needed for most other coals.

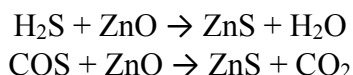
Thus, unlike in Case 5A, the syngas scrubber is eliminated in Case 5B as the syngas leaving the raw syngas coolers at around 750°F enters the HCl guard bed followed by RTI WDP directly without further quench cooling.

Warm Syngas Cleanup Process

RTI's advanced warm syngas cleanup process consists of five major system components: RTI warm gas desulfurization process (WDP), RTI direct sulfur recovery process (DSRP), RTI advanced fixed-bed water gas shift process (AFWGS), low temperature gas cooling (LTGC) and the AACRP unit. These are described in greater detail below.

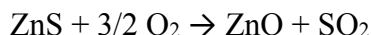
Warm Gas Desulfurization Process

The WDP process, shown in Figure 5-3, uses transport-bed reactors that are similar to commercial FCC reactor designs. It consists of a pair of these reactors: an adsorber and a regenerator. Hot syngas leaving the candle filter is routed to the WDP adsorber where it is contacted with circulating ZnO containing attrition-resistant sorbent (developed and patented by RTI) to remove the sulfur-bearing compounds, in the form of H₂S and COS, from the syngas. No COS hydrolysis pre-step is needed. The following reactions take place when the sorbent contacts the raw syngas:



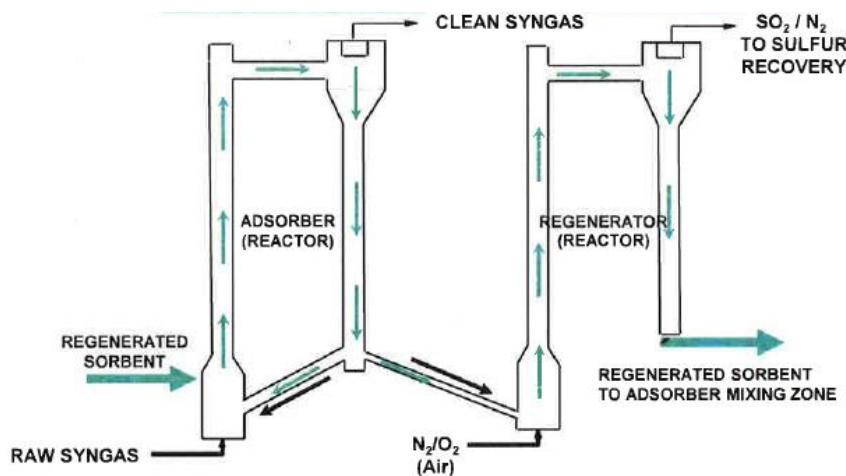
Regenerated sorbent from the regenerator, along with recycled sorbent from the adsorber standpipe, contact the raw syngas, which enters the adsorber near the bottom of the unit. The treated, essentially sulfur-free syngas is separated from sorbent via a cyclone. Any remnant attrited or fine particulate solids entrained in the essentially sulfur-free syngas are removed in a downstream filter. A majority of the sorbent separated by the cyclone is recycled to the adsorber via a standpipe, while a portion of the sorbent is fed to the regenerator.

Within the regenerator, oxidation of the ZnS containing sorbent takes place, producing SO₂ and regenerating ZnO, per the following reaction:



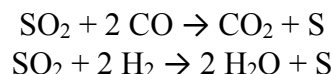
The regenerator uses air as the oxidant. Air is compressed in a multi-stage air compressor up to the regenerator operating pressure before it is fed into the regenerator. The oxidation reaction is exothermic, raising the temperature of the resulting mixture. The regenerator off-gas containing SO₂ is heat exchanged with the compressed air before the offgas enters the DSRP. The regenerated ZnO sorbent is recycled back to the adsorber, where it absorbs H₂S and COS again.

Figure 5-1
WDP Process Schematic



Direct Sulfur Recovery Process

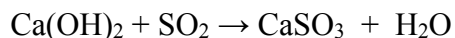
The offgas from the regenerator contains essentially SO_2 and N_2 . It goes through a downstream filter to remove any entrained solids and is cooled before it is sent into the fixed-bed DSRP reactor where SO_2 is reduced to elemental sulfur according to the following reactions:

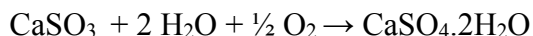


The reducing gas is provided by a hydrogen-rich syngas stream, which is obtained from pressure swing adsorption process used for hydrogen recovery from the syngas exiting the FT process. A slight excess of the reducing gas is required to ensure complete reduction of the SO_2 . Some H_2S and COS are formed alongside the elemental sulfur. The product stream from the DSRP reactor is sent onto a sulfur condenser unit where the elemental sulfur is condensed and separated. Heat is recovered in the condenser by making low pressure steam.

The condenser overhead gas still contains some residual sulfur, H_2S and COS . These are re-oxidized to SO_2 in a fixed-bed oxidation reactor containing a redox catalyst in the presence of compressed air, which functions as the oxidant. The compressed air is a slipstream drawn from the air compressor in the WDP section.

Finally, the SO_2 -containing gas leaving the oxidation reactor is cooled and sent to a lime scrubber downstream. Lime, or calcium hydroxide ($\text{Ca}(\text{OH})_2$), reacts with the SO_2 in the presence of oxygen to form gypsum ($\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$) per the following reaction, which is akin to the Wet Flue Gas Desulfurization (WFGD) process used for scrubbing flue gas.





The gypsum leaves the bottom of the scrubber as a byproduct of the CTL plant. The treated, sulfur-free overhead gas is split into three portions. One portion is recycled to the DSRP to help control reactor temperature rise due to the exothermic heat from the DSRP reactions. The second portion is compressed and routed to the WDP process to be utilized as stripping or fluidizing gas in the WDP adsorber and regenerator. The remaining portion is vented to the atmosphere.

Water Gas Shift

The treated syngas from the WDP enters RTI's advanced fixed-bed water gas shift (AFWGS) process. To achieve a 2:1 ratio of H₂ to CO in the final syngas to the F-T synthesis reactors, approximately 64 percent of the coal-derived syngas is shifted, with the remainder bypassing the shift reactors. Three shift reactors in series are used to achieve the desired H₂/CO composition in the syngas. Steam for AFWGS is provided partly by vaporized quench water in the gasifier. Additional steam is injected to the syngas stream entering the WGS reactors to push the equilibrium towards a high conversion of CO.

Low Temperature Gas Cooling

The shifted syngas goes through a series of heat exchangers to generate various levels of steam and preheat BFW, before it finally undergoes cooling by cooling water. The cooled syngas is sent to a knockout (KO) drum where the condensate is drained. Due to the absence of the wet scrubbing upstream of the RTI's WDP process, no sour water stripping is required in Case 5B.

Activated Amine CO₂ Removal Process

The AACRP is based on the activated methyldiethanolamine process, which is marketed commercially by companies such as BASF, Shell and UOP for the removal of acid gases like hydrogen sulfide (H₂S) and carbon dioxide (CO₂). The scrubbing agent is an aqueous alkaline amine solution. For this application, the CO₂-containing syngas is passed through an absorber that contains a circulating alkaline amine scrubbing solution, where the bulk of the CO₂ removal takes place. An acid-base reaction occurs where the CO₂ reacts with alkaline amine and is captured in solution.

To achieve the desired extent of CO₂ capture, and to ensure that the product gas meets the volatile organic compound (VOC) specifications, a lean amine absorber is used to treat the syngas leaving the main alkaline amine absorber. Water enters at the top of this absorber where it contacts and scrubs the CO₂-lean syngas. This serves to remove any entrained alkaline amine

The rich, CO₂-loaded solution is sent to a regenerator to release the absorbed CO₂. The solution is first depressurized, flashing off some CO₂ in the process and helping to reduce the overall heat of CO₂ regeneration. The low-pressure solution is then sent to a thermal regenerator, where heat is applied to release the remaining CO₂. Regenerated alkaline amine solution is recycled to the absorber and used again.

For the AACRP, the CO concentration in the CO₂ product is about 259 ppmv. Although this is higher than the CO limit (35 ppmv) for saline reservoir CO₂ sequestration as shown in Table 2-5, it is still within the range stated in literature (10-5,000 ppmv). It should also be noted that the CO concentration of the AACRP CO₂ product is one order of magnitude lower than in Case 5A as AACRP is more selective towards absorbing CO₂ when compared with the Rectisol® process.

Fischer Tropsch Synthesis and Product Upgrade

FT synthesis and product upgrade for the Case 5B plant is very similar to Case 5A for the most part. Except that an additional low temperature water gas shift process is required to meet the hydrogen requirement of the DSRP process. Hence, the syngas exiting the hydrocarbon recovery block is shifted to achieve about 90% CO conversion. This not only results in some additional capital cost for the WGS reactors and the PSA system for hydrogen recovery, but also requires steam which increases the parasitic load. The other downside to this WGS step is the reduced energy content in the fuel gas going to the power block.

Figure 5-2
Case 5B Plant – Overall BFD

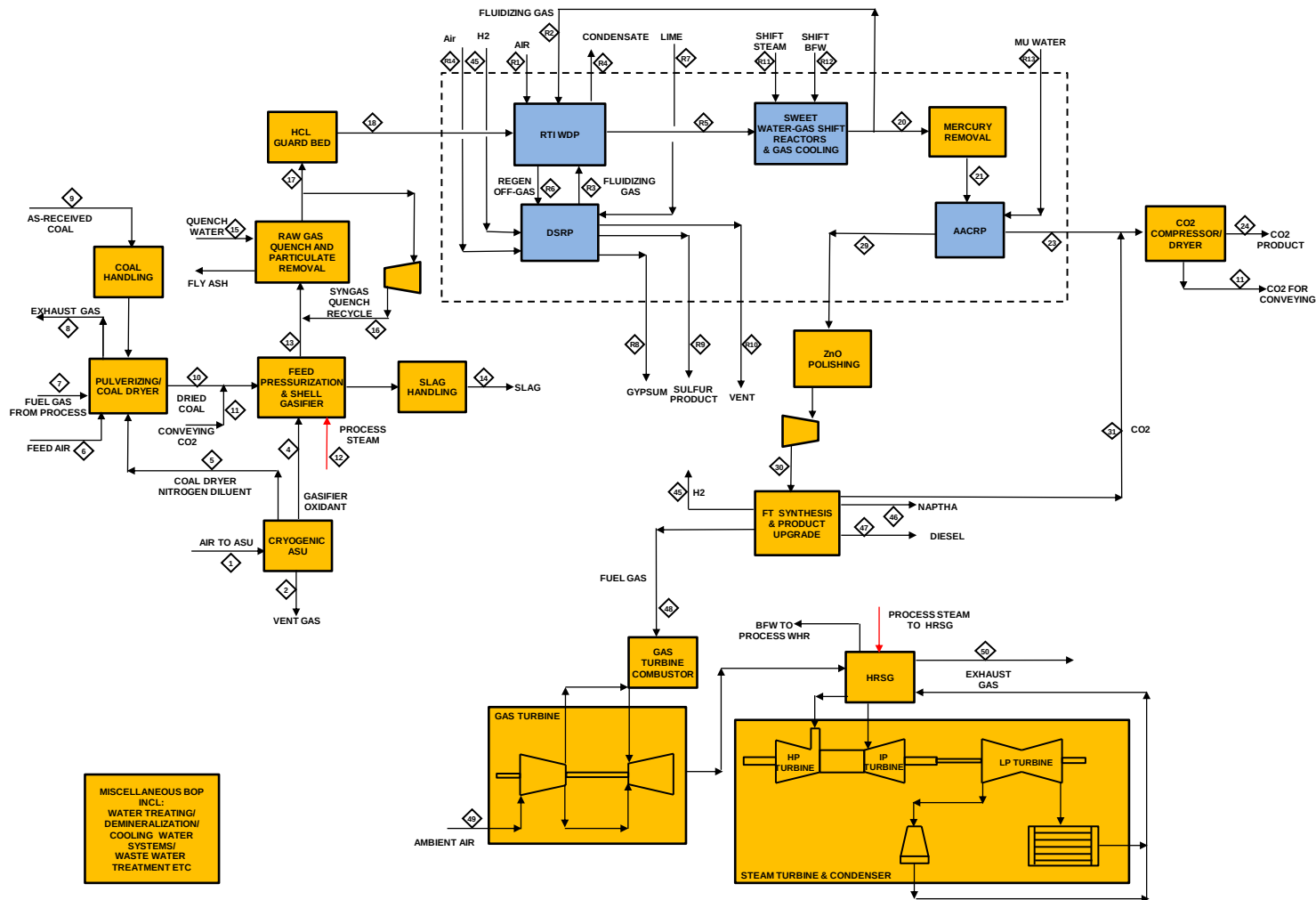


Table 5-1
Case 5B Plant – Overall Stream Table

	1	2	4	5	6	7	8	9	10	11	12	13
	Air to ASU	Vent Gas from ASU	Gasifier Oxidant	Coal Dryer Nitrogen Diluent	Air to Coal Dryer	Fuel Gas to Coal Dryer	Exhaust Gas from Coal Dryer	As-Received Coal	Dried Coal	Conveying CO2	Gasifier Steam	Raw Syngas
Mole Frac												
H2	0.0000	0.0000	0.0000	0.0000	0.0000	0.1267	0.0000	0.0000	0.0000	0.0000	0.0000	0.2894
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0168	0.0000	0.0000	0.0000	0.0000	0.0000	0.6093
CO2	0.0003	0.0077	0.0000	0.0000	0.0003	0.1976	0.0114	0.0000	0.0000	1.0000	0.0000	0.0263
CH4	0.0000	0.0000	0.0000	0.0000	0.0000	0.1592	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003
H2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0082
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0007
NH3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0010
HCL	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0009
N2	0.7732	0.7077	0.0178	0.9919	0.7732	0.2246	0.8926	0.0000	0.0000	0.0000	0.0000	0.0093
AR	0.0092	0.0225	0.0318	0.0023	0.0092	0.1503	0.0051	0.0000	0.0000	0.0000	0.0000	0.0084
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0099	0.0714	0.0000	0.0004	0.0099	0.0049	0.0791	0.0000	0.0000	0.0000	1.0000	0.0460
O2	0.2074	0.1906	0.9504	0.0054	0.2074	0.0000	0.0118	0.0000	0.0000	0.0000	0.0000	0.0000
SO2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Cl2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0027	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0212	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0202	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0259	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0174	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0119	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0197	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
GYPSUM	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
LIME	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mole Flow lbmol/hr	197,257	8,354	40,525	79,317	10,589	1,457	97,683			4,970	9,376	153,005
Mass Flow lb/hr	5,692,010	237,668	1,304,140	2,225,520	305,560	43,619	2,687,470	1,750,520	1,637,750	218,718	168,908	3,154,190
Temperature F	59	89.6	241	70	59	87	180	59	60	376	650	2600
Pressure psia	15	190	940	18	15	15	15	15	840	785	740	615
Solid Frac	0	0	0	0	0	0	0	0	0	0	0	0
Enthalpy Btu/hr	-2.40E+08	-8.38E+07	4.09E+07	-7.50E+06	-1.28E+07	-	-9.21E+08	-1.58E+09	-8.08E+08	-8.29E+08	-9.38E+08	-2.85E+09
Average MW	28.856	28.451	32.181	28.058	28.856	29.930	27.512			44.010	18.015	20.615

Table 5-1 (cont'd)
Case 5B Plant – Overall Stream Table

	14	15	17	18	20	21	23	29	30	45	R1	R2
	Slag	Quench Water	Syngas to HCl Removal	Syngas to Shift	Syngas to Hg Removal	Syngas to AACRP	Total Captured CO2	Clean Syngas from AACRP	Treated Syngas to FT	Hydrogen to DSRP	Air to WDP	Fluidizing Gas to WDP
Mole Frac												
H2	0.0000	0.0000	0.1855	0.1854	0.4790	0.4790	0.0007	0.6537	0.6537	0.9900	0.0000	0.4790
CO	0.0000	0.0000	0.3904	0.3902	0.2303	0.2303	0.0002	0.3143	0.3143	0.0000	0.0000	0.2303
CO2	0.0000	0.0000	0.0169	0.0175	0.2726	0.2726	0.9108	0.0046	0.0046	0.0000	0.0003	0.2726
CH4	0.0000	0.0000	0.0002	0.0002	0.0002	0.0002	0.0000	0.0003	0.0003	0.0000	0.0000	0.0002
H2S	0.0000	0.0000	0.0053	0.0053	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS	0.0000	0.0000	0.0005	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH3	0.0000	0.0000	0.0006	0.0006	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL	0.0000	0.0000	0.0006	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N2	0.0000	0.0000	0.0060	0.0060	0.0100	0.0100	0.0000	0.0136	0.0136	0.0100	0.7732	0.0100
AR	0.0000	0.0000	0.0054	0.0054	0.0067	0.0067	0.0000	0.0091	0.0091	0.0000	0.0092	0.0067
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0000	1.0000	0.3887	0.3891	0.0012	0.0012	0.0882	0.0043	0.0043	0.0000	0.0099	0.0012
O2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2074	0.0000
SO2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Cl2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
GYPSUM	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
LIME	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mole Flow lbmol/hr		85,773	238,778	238,922	193,879	193,879	57,311	141,988	141,988	1,236	9,883	3,187
Mass Flow lb/hr	175,324	1,545,220	4,699,410	4,703,100	3,874,800	3,874,800	2,388,850	1,583,590	1,583,590	2,813	285,190	63,686
Temperature F	140	392	750	750	95	95	122	127	200	85	59	95
Pressure psia	615	685	595	585	498	488	21	496	371	390	15	498
Solid Frac		0	0	0	0	0	0	0	0		0	0
Enthalpy Btu/hr	1988530	-1.01E+10	-1.36E+10	-1.36E+10	-1.11E+10	-1.11E+10	-9.36E+09	-2.25E+09	-2.18E+09	8.31E+04	-1.20E+07	-1.82E+08
Average MW		18.015	19.681	19.685	19.986	19.986	41.682	11.153	11.153	2.276	28.856	19.986

Table 5-1 (cont'd)
Case 5B Plant – Overall Stream Table

	R3	R4	R5	R6	R7	R8	R9	R10	R11	R12	R13	R14
	Stripping Gas to WDP	Condensate	Syngas to Shift	WDP Regenerator Off-Gas	Lime	Gypsum	Sulfur	Vent	Shift Steam	Shift BFW	MU Water	Air to DSRP
Mole Frac												
H2	0.0000	0.0000	0.1888	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.3873	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	0.0000	0.0000	0.0212	0.0003	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003
CH4	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH3	0.0000	0.0000	0.0006	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N2	0.9810	0.0000	0.0081	0.8475	0.0000	0.0002	0.0000	0.9810	0.0000	0.0000	0.0000	0.7788
AR	0.0116	0.0000	0.0054	0.0101	0.0000	0.0000	0.0000	0.0116	0.0000	0.0000	0.0000	0.0093
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0074	1.0000	0.3883	0.0033	0.9737	0.9788	0.0000	0.0074	1.0000	1.0000	1.0000	0.0028
O2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2089
SO2	0.0000	0.0000	0.0000	0.1388	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Cl2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000
GYPSUM	0.0000	0.0000	0.0000	0.0000	0.0000	0.019882	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
LIME	0.0000	0.0000	0.0000	0.0000	0.0263	0.000895	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mole Flow lbmol/hr	1,228	71	242,622	9,844	5,022	6,359	557	4,085	14,169	6,798	5,420	548
Mass Flow lb/hr	34,486	1,274	4,759,240	325,941	97,887	129,741	17,855	114,698	255,266	122,474	97,643	15,857
Temperature F	172	95	680	888	75	214	291	126	550	490	122	269
Pressure psia	415	136	563	556	15	15	15	15	615	650	535	589
Solid Frac	0	0	0	0	0	0	0	0	0	0	0	0
Enthalpy Btu/hr	-2.12E+05	-8.72E+06	-1.41E+10	-1.17E+08	-6.57E+08	-8.12E+08	1.24E+06	-1.77E+06	-1.43E+09	-7.83E+08	-6.66E+08	4.95E+05
Average MW	28.079	18.015	19.616	33.110	19.490	20.105	245.184	28.078	18.015	18.015	18.015	28.934

Table 5-2
Case 5B Plant –FT Process Stream Table

	30	31	32	33	34	35	36	37	38	39
Component Mole Fraction										
O2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2	0.01361	0.16943	0.04569	0.16943	0.00384	0.00000	0.17933	0.17995	0.16943	0.00000
CO	0.31434	0.15931	0.28242	0.15931	0.00588	0.00000	0.16861	0.16920	0.15931	0.00000
CO2	0.00461	0.05546	0.01508	0.05546	0.00912	1.00000	0.00029	0.00029	0.05546	0.00000
H2	0.65373	0.32574	0.58621	0.32574	0.00490	0.00000	0.34477	0.34597	0.32574	0.00000
H2O	0.00427	0.00327	0.00407	0.00327	0.06074	0.00000	0.00346	0.00000	0.00327	0.00000
NH3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCN	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H2S	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
AR	0.00911	0.11309	0.03052	0.11309	0.00395	0.00000	0.11970	0.12011	0.11309	0.00000
CH4	0.00032	0.11576	0.02409	0.11576	0.00617	0.00000	0.12252	0.12295	0.11576	0.00000
C2H4	0.00000	0.00199	0.00041	0.00199	0.00034	0.00000	0.00211	0.00211	0.00199	0.00000
C2H6	0.00000	0.01028	0.00212	0.01028	0.00246	0.00000	0.01088	0.01092	0.01028	0.00000
C3H6	0.00000	0.01335	0.00275	0.01335	0.00845	0.00000	0.01413	0.01418	0.01335	0.00000
C3H8	0.00000	0.00603	0.00124	0.00603	0.00435	0.00000	0.00638	0.00641	0.00603	0.00000
IC4H8	0.00000	0.00046	0.00010	0.00046	0.00085	0.00000	0.00049	0.00049	0.00046	0.00000
C4H8	0.00000	0.00921	0.00190	0.00921	0.01723	0.00000	0.00975	0.00978	0.00921	0.00000
IC4H10	0.00000	0.00022	0.00005	0.00022	0.00036	0.00000	0.00023	0.00023	0.00022	0.00000
C4H10	0.00000	0.00403	0.00083	0.00403	0.00880	0.00000	0.00427	0.00428	0.00403	0.00000
C5H10	0.00000	0.00262	0.00054	0.00262	0.01414	0.00000	0.00277	0.00278	0.00262	0.06619
NC5H12	0.00000	0.00445	0.00092	0.00445	0.02900	0.00000	0.00471	0.00473	0.00445	0.12547
IC5H12	0.00000	0.00058	0.00012	0.00058	0.00287	0.00000	0.00061	0.00061	0.00058	0.01393
C6H12	0.00000	0.00114	0.00023	0.00114	0.01811	0.00000	0.00121	0.00121	0.00114	0.06000
NC6H14	0.00000	0.00182	0.00038	0.00182	0.03578	0.00000	0.00193	0.00193	0.00182	0.11373
IC6H14	0.00000	0.00026	0.00005	0.00026	0.00373	0.00000	0.00027	0.00027	0.00026	0.01263
C7H14	0.00000	0.00038	0.00008	0.00038	0.01919	0.00000	0.00040	0.00040	0.00038	0.05434
C7H16	0.00000	0.00068	0.00014	0.00068	0.04097	0.00000	0.00071	0.00072	0.00068	0.11442
C8H16	0.00000	0.00011	0.00002	0.00011	0.01839	0.00000	0.00012	0.00012	0.00011	0.04919
C8H18	0.00000	0.00019	0.00004	0.00019	0.03888	0.00000	0.00020	0.00021	0.00019	0.10357
C9H18	0.00000	0.00003	0.00001	0.00003	0.01696	0.00000	0.00003	0.00003	0.00003	0.04456
C9H20	0.00000	0.00005	0.00001	0.00005	0.03575	0.00000	0.00005	0.00005	0.00005	0.09382
C10H20	0.00000	0.00001	0.00000	0.00001	0.01545	0.00000	0.00001	0.00001	0.00001	0.04038
C10H22	0.00000	0.00001	0.00000	0.00001	0.03253	0.00000	0.00002	0.00002	0.00001	0.08502
C11H22	0.00000	0.00000	0.00000	0.00000	0.00144	0.00000	0.00000	0.00000	0.00000	0.00000

Table 5-2 (cont'd)
Case 5B Reference Plant – FT Process Stream Table

	30	31	32	33	34	35	36	37	38	39
Component Mole										
C11H24	0.00000	0.00001	0.00000	0.00001	0.04208	0.00000	0.00001	0.00001	0.00001	0.00000
C12H24	0.00000	0.00000	0.00000	0.00000	0.00130	0.00000	0.00000	0.00000	0.00000	0.00000
C12H26	0.00000	0.00000	0.00000	0.00000	0.03813	0.00000	0.00000	0.00000	0.00000	0.00000
C13H26	0.00000	0.00000	0.00000	0.00000	0.00118	0.00000	0.00000	0.00000	0.00000	0.00000
C13H28	0.00000	0.00000	0.00000	0.00000	0.03456	0.00000	0.00000	0.00000	0.00000	0.00000
C14H28	0.00000	0.00000	0.00000	0.00000	0.00107	0.00000	0.00000	0.00000	0.00000	0.00000
C14H30	0.00000	0.00000	0.00000	0.00000	0.03131	0.00000	0.00000	0.00000	0.00000	0.00000
C15H30	0.00000	0.00000	0.00000	0.00000	0.00097	0.00000	0.00000	0.00000	0.00000	0.00000
C15H32	0.00000	0.00000	0.00000	0.00000	0.02836	0.00000	0.00000	0.00000	0.00000	0.00000
C16H32	0.00000	0.00000	0.00000	0.00000	0.00088	0.00000	0.00000	0.00000	0.00000	0.00000
C16H34	0.00000	0.00000	0.00000	0.00000	0.02569	0.00000	0.00000	0.00000	0.00000	0.00000
C17H34	0.00000	0.00000	0.00000	0.00000	0.00079	0.00000	0.00000	0.00000	0.00000	0.00000
C17H36	0.00000	0.00000	0.00000	0.00000	0.02327	0.00000	0.00000	0.00000	0.00000	0.00000
C18H36	0.00000	0.00000	0.00000	0.00000	0.00072	0.00000	0.00000	0.00000	0.00000	0.00000
C18H38	0.00000	0.00000	0.00000	0.00000	0.02108	0.00000	0.00000	0.00000	0.00000	0.00000
C19H38	0.00000	0.00000	0.00000	0.00000	0.00065	0.00000	0.00000	0.00000	0.00000	0.00000
C19H40	0.00000	0.00000	0.00000	0.00000	0.01909	0.00000	0.00000	0.00000	0.00000	0.00000
C20H40	0.00000	0.00000	0.00000	0.00000	0.00059	0.00000	0.00000	0.00000	0.00000	0.00000
C20H42	0.00000	0.00000	0.00000	0.00000	0.01730	0.00000	0.00000	0.00000	0.00000	0.00000
C21OP	0.00000	0.00000	0.00000	0.00000	0.01658	0.00000	0.00000	0.00000	0.00000	0.00000
C22OP	0.00000	0.00000	0.00000	0.00000	0.01537	0.00000	0.00000	0.00000	0.00000	0.00000
C23OP	0.00000	0.00000	0.00000	0.00000	0.01424	0.00000	0.00000	0.00000	0.00000	0.00000
C24OP	0.00000	0.00000	0.00000	0.00000	0.01320	0.00000	0.00000	0.00000	0.00000	0.00000
C25OP	0.00000	0.00000	0.00000	0.00000	0.01224	0.00000	0.00000	0.00000	0.00000	0.00000
C26OP	0.00000	0.00000	0.00000	0.00000	0.01134	0.00000	0.00000	0.00000	0.00000	0.00000
C27OP	0.00000	0.00000	0.00000	0.00000	0.01051	0.00000	0.00000	0.00000	0.00000	0.00000
C28OP	0.00000	0.00000	0.00000	0.00000	0.00974	0.00000	0.00000	0.00000	0.00000	0.00000
C29OP	0.00000	0.00000	0.00000	0.00000	0.00903	0.00000	0.00000	0.00000	0.00000	0.00000
C30WAX	0.00000	0.00000	0.00000	0.00000	0.11746	0.00000	0.00000	0.00000	0.00000	0.00000
OXVAP	0.00000	0.00000	0.00000	0.00000	0.00255	0.00000	0.00000	0.00000	0.00000	0.00000
OXHC	0.00000	0.00002	0.00000	0.00002	0.01808	0.00000	0.00002	0.00002	0.00002	0.02274
OXH2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
GASOLINE	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
DIESEL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Mole Flow (LBMOL/HR)	141,988	36,811	178,799	48,132	2,637	625	10,696	10,659	36,811	1,011
Mass Flow (LB/HR)	1,583,590	798,816	2,382,410	1,044,480	558,147	27,495	218,166	217,498	798,816	103,761
Temperature (F)	200	149	188	100	286	100	100	100	100	258
Pressure (PSIA)	371	378	371	284	75	275	275	430	284	50

Table 5-2 (cont'd)
Case 5B Reference Plant – FT Process Stream Table

	40	41	42	43	44	45	46	47	51	52
Component Mole Fraction										
O2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.16360
CO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00976
CO2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.14377
H2	0.00000	0.00000	1.00000	1.00000	1.00000	1.00000	0.00000	0.00000	0.00000	0.45954
H2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.00000	0.00153
NH3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCN	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H2S	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
AR	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.10790
CH4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.10768
C2H4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00114
C2H6	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00387
C3H6	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00084
C3H8	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00030
IC4H8	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H8	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00005
IC4H10	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H10	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00002
C5H10	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NC5H12	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.07035	0.00000	0.00000	0.00000
IC5H12	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.03069	0.00000	0.00000	0.00000
C6H12	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NC6H14	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.15203	0.00000	0.00000	0.00000
IC6H14	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.05373	0.00000	0.00000	0.00000
C7H14	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H16	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.43985	0.00000	0.00000	0.00000
C8H16	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H18	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.06049	0.00000	0.00000	0.00000
C9H18	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H20	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H20	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H22	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C11H22	0.00509	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000

Table 5-2 (cont'd)
Case 5B Reference Plant – FT Process Stream Table

	40	41	42	43	44	45	46	47	51	52
Component Mole Fraction										
C11H24	0.14928	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C12H24	0.00461	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C12H26	0.13522	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C13H26	0.00418	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C13H28	0.12254	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C14H28	0.00379	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C14H30	0.11100	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C15H30	0.00343	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C15H32	0.10056	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C16H32	0.00311	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C16H34	0.09108	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C17H34	0.00282	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C17H36	0.08251	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C18H36	0.00255	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C18H38	0.07473	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C19H38	0.00231	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C19H40	0.06769	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H40	0.00000	0.00238	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H42	0.00000	0.06985	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C21OP	0.00000	0.06696	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C22OP	0.00000	0.06207	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C23OP	0.00000	0.05753	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C24OP	0.00000	0.05333	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C25OP	0.00000	0.04942	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C26OP	0.00000	0.04580	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C27OP	0.00000	0.04245	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C28OP	0.00000	0.03934	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C29OP	0.00000	0.03646	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C30WAX	0.00000	0.47439	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
OXVAP	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
OXHC	0.03348	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
OXH2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
GASOLINE	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.19286	0.00000	0.00000	0.00000
DIESEL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.00000	0.00000	0.00000
Mole Flow (LBMOL/HR)	744	653	502	40	1,273	2,473	1,490	1,832	5,433	11,662
Mass Flow (LB/HR)	148,764	305,240	1,011	81	2,566	4,985	150,136	389,211	97,877	214,311
Temperature (F)	258	258	100	100	602	85	100	100	650	85
Pressure (PSIA)	50	50	50	50	50	385	50	50	615	385

5.2 SPARING PHILOSOPHY

The sparing philosophy for the RTI-provided equipment for the Case 5B plant is provided below. The design has:

- Three WDP trains (3 x 33%)
- Three AACRP trains (3 x 33%)
- Two DSRP trains (2 x 50%)
- Two AFWGS trains (2 x 50%)

5.3 PERFORMANCE RESULTS

Similar to Case 5A, Case 5B plant consumes 21,006 tpd of Illinois #6 coal at the Midwestern site and produces 50,000 bbl/day of FT liquid product, containing 70% FT diesel and 30% FT Naphtha. Overall performance for the Case 5B plant is summarized in Table 5-3, which includes auxiliary power requirements. Loads that are unique to the RTI WDP system are shown in bold and italicized.

Table 5-3
Case 5B Plant Performance Summary

Power Cycle Power Generation	Electrical Load, kWe
Gas Turbine Power	101,455
Steam Turbine Power	448,585
Syngas Turboexpansion Power	9,995
Total	560,035
CTL Auxiliary Load Description	Electrical Load, kWe
Coal Handling	1,760
Coal Milling	8,160
Slag Handling	2,080
ASU Auxiliaries	3,739
ASU Main Air Compressor	220,132
Oxygen Compressor	36,324
Quench Water Pump	2,223
Syngas Recycle Compressor	2,920
RTI WDP	20,730
AACRP	39,464
DSRP	3,083
CO ₂ Compressor Auxiliary	73,449
FT Power Requirement	26,824
Miscellaneous BOP	13,908
Circulating Water Pumps	14,463
Ground Water Pumps	800
Cooling Tower Fans	7,572
Gas Turbine Auxiliaries	4,000
Steam Turbine Auxiliaries	400
Transformer Losses	3,304
Condensate Pumps	200
Boiler Feed Water Pumps	4,335
Total	489,869
NET PLANT POWER, kWe	70,166

Table 5-4 shows the carbon balance for the Case 5B plant. Carbon leaves the coal-to-FT liquids process plant as unburned carbon in the sour water and LTGC condensate, slag, CO₂ in the dryer exhaust gas, ASU vent gas, FT products, sulfur products, and both the AACRP CO₂ capture products. Additionally, carbon also leaves the power cycle as CO₂ in the stack. Carbon in the FT products is considered to be product and not emissions. The carbon capture efficiency is defined

as the amount of carbon in the product streams, which include the FT liquids and captured CO₂ products from the Rectisol® AGR and the post FT AACRP process, relative to the amount of carbon in the coal feedstock less carbon contained in the slag. For Case 5B, the carbon capture efficiency is 92.8%.

Table 5-4
Case 5B Plant – Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	1,115,957	
ASU Air	774	
Carbon in for HCl Removal (as NaHCO ₃)	1,736	
Coal Dryer Air	38	
Carbon in Air to WDP	36	
Carbon in Air to DSRP		
Carbon in Sour water		2,752
Carbon in LTGC condensate		87
ASU Vent		774
Exhaust Gas to Stack		78,359
CO ₂ (post FT)		7,503
Carbon in Slag		5,579
Sulfur Product		
Gypsum		38
DSRP Vent		
FT Naphtha		125,976
FT Diesel		330,084
AACRP CO ₂ Product		567,388
Total	1,118,541	1,118,541

Table 5-5 shows the sulfur balance for the Case 5B plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the elemental sulfur recovered in the DSRP, sulfur recovered in the form of gypsum leaving the lime scrubber, sulfur emitted in the coal dryer exhaust gas, sulfur emitted in the power cycle exhaust gas, sulfur emitted in the FT liquids product, sulfur emitted in the DSRP vent gas and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible. There is no net sulfur emission from the Case 5B CTL plant.

Table 5-5
Case 5B Plant – Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	43,855	
Coal Dryer Exhaust Gas		-
GT Exhaust Gas		-
CO ₂ (post FT)		-

Sulfur Product		34,324
Gypsum		9,492
DSRP Vent		-
FT Naphtha		-
FT Diesel		-
AACRP CO ₂ Product		39
Total	43,855	43,855

Table 5-6 shows the overall water balance for the Case 5B plant. Raw water withdrawal is the amount of raw water consumed by the plant. Some water is internally recycled and consumed by the slag handling process, syngas quench process and as makeup to the cooling tower.

Table 5-6 Case 5B Plant – Overall Water Balance

Water Usage by Area	Water Demand		Internal Recycle		Raw water Withdrawal		Process Water Discharge		Raw Water Consumption	
	m ³ /min	gpm	m ³ /min	gpm	m ³ /min	gpm	m ³ /min	gpm	m ³ /min	gpm
Slag Handling	1.7	456	1.7	456	-	-	-	-	-	-
Quench/Wash	11.7	3,084	7.6	2,001	4.1	1,083	-	-	4.1	1,083
AGR Makeup Water	0.7	195	-	-	0.7	195	-	-	0.7	195
SWS Blowdown	-	-	-	-	-	-	0.1	30	(0.1)	(30)
FT-Area	0.1	37	-	-	0.1	37	-	-	0.1	37
Condenser Makeup	5.3	1,411		-	5.3	1,411		-	5.3	1,411
Gasifier Steam	1.3	337	-	-	1.3	337	-	-	1.3	337
Shift Steam	3.6	949	-	-	3.6	949	-	-	3.6	949
BFW Makeup	0.5	125	-	-	0.5	125	-	-	0.5	125
Cooling Tower	57.0	15,053	-	-	57.0	15,053	14.2	3,756	42.8	11,297
BFW Blowdown	-	-	0.5	125	(0.5)	(125)	-	-	(0.5)	(125)
SWS Blowdown	-	-	1.1	302	(1.1)	(302)	-	-	(1.1)	(302)
TOTAL	76.6	20,237		2,883	65.7	17,354		3,786	51.4	13,567

5.4 CAPITAL COST

Table 5-7 shows the cost breakdown of the Case 5B plant, consistent with the Code of Accounts format as expressed in the CTL Study. The accounts/subaccounts of interest for this study are:

- 4 Gasifier & Accessories,
- 5A Gas Cleanup & Piping and
- 5AA FT Synthesis and Product Upgrade
- 5B CO₂ Removal and Compression

These are shown with more detail to include the various subaccounts and to provide more clarity as to wherein the major cost differences lie among the cases.

Table 5-8 shows the calculation and addition of owner's costs to determine the TOC, which is used to calculate the cost of production for FT products.

The estimated TOC of the Case 5B plant in 2011 dollars is \$6,530 MM.

Table 5-7
Case 5B Total Plant Cost Summary

Case 5b: Shell Gasifier w/RTI WDP CFTL Plant				Coal Feed, lb/hr 1,750,518		Plant Size		50,000 barrels per day of FT-liquids				
Illinois #6				Coal HHV, Btu/lb 13,126 (dry basis)		Cost Base		Jun 2011 (\$x1000)				
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost \$	Eng'g CM H.O. & Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect				Process	Project	\$	
1	COAL & SORBENT HANDLING	\$38,319	\$6,719	\$29,308	\$0	\$0	\$74,346	\$6,589	\$0	\$16,187	\$97,122	
2	COAL & SORBENT PREP & FEED											
2.3	Dry Coal Injection System	\$189,497	\$2,184	\$17,357	\$0	\$0	\$209,038	\$17,410	\$0	\$45,290	\$271,738	
2.x	Other Coal & Sorbent Prep & Feed Systems	\$130,490	\$22,641	\$35,142	\$0	\$0	\$188,273	\$15,985	\$0	\$40,851	\$245,109	
	SUBTOTAL 2.	\$319,987	\$24,825	\$52,499	\$0	\$0	\$397,311	\$33,395	\$0	\$86,141	\$516,847	
3	FEEDWATER & MISC BOP SYSTEMS	\$13,426	\$9,083	\$14,140	\$0	\$0	\$36,649	\$3,395	\$0	\$9,770	\$49,814	
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries	\$699,517	\$0	\$301,659	\$0	\$0	\$1,001,176	\$88,113	\$137,204	\$188,698	\$1,415,191	
4.3	ASU/Oxidant Compression	\$525,869	\$0 w/equip.	\$0	\$0	\$0	\$525,869	\$49,137	\$0	\$57,501	\$632,507	
4.4	LT Heat Recovery & FG Saturation	\$82,362	\$0	\$26,146	\$0	\$0	\$108,508	\$10,252	\$0	\$23,751	\$142,511	
4.6	Flare Stack System	\$0	\$3,422	\$1,382	\$0	\$0	\$4,804	\$449	\$0	\$1,050	\$6,303	
4.9	Gasification Foundations	\$0	\$41,427	\$24,706	\$0	\$0	\$66,133	\$6,096	\$0	\$18,058	\$90,287	
	SUBTOTAL 4.	\$1,307,747	\$44,849	\$353,894	\$0	\$0	\$1,706,490	\$154,047	\$137,204	\$289,058	\$2,286,799	
5A	GAS CLEANUP & PIPING											
5A.1a	RTI WDP for Sulfur Removal	\$107,717	\$0	\$129,261	\$0	\$0	\$236,978	\$22,360	\$47,395	\$61,347	\$368,080	
5A.1b	AA CO2 Removal Process	\$73,255	\$0	\$59,527	\$0	\$0	\$132,782	\$12,529	\$26,556	\$34,373	\$206,240	
5A.2	Elemental Sulfur Plant	\$23,585	\$0	\$25,291	\$0	\$0	\$48,876	\$4,642	\$9,775	\$12,659	\$75,953	
5A.3	Mercury Removal	\$2,985	\$0	\$2,256	\$0	\$0	\$5,241	\$494	\$262	\$1,199	\$7,196	
5A.4	Shift Reactors	\$4,822	\$0	\$1,929	\$0	\$0	\$6,751	\$630	\$0	\$1,476	\$8,858	
5A.5	Particulate Removal	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
5A.6	Blowback Gas Systems	\$3,633	\$611	\$342	\$0	\$0	\$4,586	\$423	\$0	\$1,002	\$6,011	
5A.7	Fuel Gas Piping	\$0	\$2,074	\$1,357	\$0	\$0	\$3,431	\$299	\$0	\$747	\$4,477	
5A.9	HGCU Foundations	\$0	\$1,965	\$1,324	\$0	\$0	\$3,289	\$304	\$0	\$1,078	\$4,671	
5A.10	Syngas Conditioning	\$26,360	\$18,873	\$22,419	\$0	\$0	\$67,652	\$6,427	\$0	\$14,816	\$88,894	
5A.11	HCl Guard Bed	\$5,886	\$2,633	\$3,408	\$0	\$0	\$11,928	\$1,133	\$0	\$2,612	\$15,673	
	SUBTOTAL 5.	\$248,244	\$26,156	\$247,114	\$0	\$0	\$521,514	\$49,241	\$83,989	\$131,309	\$786,053	
5AA	F-T SYNTHESIS AND PRODUCT UPGRADE											
5AA.1	FT Synthesis*	\$141,029	\$0	\$0	\$0	\$0	\$141,029	\$13,539	\$38,078	\$48,162	\$240,807	
5AA.2	Amine CO2 Absorption	\$6,497 w/equip.	w/equip.	\$0	\$0	\$0	\$6,497	\$613	\$1,299	\$1,682	\$10,091	
5AA.4	Overhead Recovery	\$8,255	\$0	\$0	\$0	\$0	\$8,255	\$793	\$2,229	\$2,819	\$14,096	
5AA.5	Product Separation	\$35,082	\$0	\$0	\$0	\$0	\$35,082	\$3,368	\$9,472	\$11,980	\$59,902	
5AA.6	Hydrogen Recovery	\$11,172	\$0	\$0	\$0	\$0	\$11,172	\$1,073	\$3,016	\$3,815	\$19,075	
5AA.8	Naphtha Hydrotreater	\$13,247	\$0	\$0	\$0	\$0	\$13,247	\$1,271	\$3,577	\$4,524	\$22,619	
5AA.9	Diesel Hydrotreater	\$33,270	\$0	\$0	\$0	\$0	\$33,270	\$3,195	\$8,983	\$11,362	\$56,809	
5AA.10	Wax Hydrocracker	\$81,822	\$0	\$0	\$0	\$0	\$81,822	\$7,855	\$22,093	\$27,942	\$139,712	
5AA.11	Raw Fuel Gas Compressor, HP	\$25,916 w/equip.	w/equip.	\$0	\$0	\$0	\$25,916	\$2,417	\$0	\$5,667	\$34,000	
5AA.12	Fuel Gas to GT Compressor, HP	\$6,830 w/equip.	w/equip.	\$0	\$0	\$0	\$6,830	\$637	\$0	\$1,493	\$8,960	
	SUBTOTAL 5AA.	\$363,118	\$0	\$0	\$0	\$0	\$363,118	\$34,760	\$88,747	\$119,447	\$606,072	
5B	CO2 Compression & Drying	\$83,268	\$0	\$31,391	\$0	\$0	\$114,659	\$10,692	\$0	\$25,071	\$150,422	
	SUBTOTAL 5B.	\$83,268	\$0	\$31,391	\$0	\$0	\$114,659	\$10,692	\$0	\$25,071	\$150,422	
6	COMBUSTION TURBINE/ACCESSORIES	\$56,131	\$467	\$4,517	\$0	\$0	\$61,115	\$13,230	\$14,104	\$17,589	\$106,037	
7	HRSG, DUCTING & STACK	\$16,763	\$1,471	\$4,771	\$0	\$0	\$23,005	\$2,121	\$0	\$2,926	\$28,052	
8	STEAM TURBINE GENERATOR	\$69,180	\$1,562	\$20,266	\$0	\$0	\$91,009	\$7,926	\$0	\$13,211	\$112,146	
9	COOLING WATER SYSTEM	\$15,812	\$21,082	\$18,297	\$0	\$0	\$55,191	\$4,952	\$0	\$12,752	\$72,895	
10	ASH/SPENT SORBENT HANDLING SYS	\$50,895	\$3,552	\$25,045	\$0	\$0	\$79,493	\$7,430	\$0	\$9,426	\$96,349	
11	ACCESSORY ELECTRIC PLANT	\$41,256	\$24,129	\$42,081	\$0	\$0	\$107,466	\$9,109	\$0	\$23,162	\$139,737	
12	INSTRUMENTATION & CONTROL	\$15,804	\$3,203	\$10,391	\$0	\$0	\$29,398	\$2,616	\$1,470	\$5,614	\$39,099	
13	IMPROVEMENTS TO SITE	\$7,961	\$6,045	\$24,023	\$0	\$0	\$38,028	\$3,765	\$0	\$12,538	\$54,331	
14	BUILDINGS & STRUCTURES	\$0	\$18,139	\$17,521	\$0	\$0	\$35,660	\$3,169	\$0	\$6,200	\$45,029	
	CALCULATED TOTAL COST	\$2,647,912	\$191,282	\$895,258	\$0	\$0	\$3,734,452	\$346,437	\$325,513	\$780,402	\$5,186,803	

Table 5-8
Case 5B Total Overnight Cost Breakdown

Owner's Costs	\$ x \$1,000
Preproduction Costs	
6 months All Labor	\$47,584
1 Month Maintenance Materials	\$8,116
1 Month Non-Fuel Consumables	\$6,099
1 Month Waste Disposal	\$1,617
25% of 1 Months Fuel Cost at 100% CF	\$10,958
2% of TPC	\$103,736
Total	\$178,110
Inventory Capital	
60-day supply of fuel at 100% CF	\$86,461
60-day supply of non-fuel consumables at 100% CF	\$12,031
0.5% of TPC (spare parts)	\$25,934
Total	\$124,426
Initial Cost for Catalyst and Chemicals	\$121,848
Land	\$900
Other Owner's Cost	\$774,948
Financing Costs	\$139,491
Total Owner's Costs	\$1,337,825
Total Overnight Costs (TOC)	\$6,530,152
TASC Multiplier	1.147
Total As-Spent Cost (TASC)	\$7,490,085

5.5 OPERATING COSTS

Table 5-9 shows the operating cost breakdown for the Case 5B Plant.

Table 5-9
Case 5B Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Shell Gasifier w/RTI WDP CFTL Plant				
Plant Size (bbl/day FT Liquids)	50,000				
Primary/Secondary Fuel:	Illinois #6 Coal				
Design/Construction	5 years	Book Life (yrs):		20	
TPC (Plant Cost) Year	June 2011	TPI Year:		2016	
Capacity Factor (%)	90	CO2 Captured (TPD)		25,280	
OPERATING & MAINTENANCE LABOR					
Operating Labor					
Operating Labor Rate (base):	\$39.70	\$/hr			
Operating Labor Burden:	30.00	% of base			
Labor Overhead Charge	25.00	% of labor			
Operating Labor Requirements per Shift	units/mod	Total Plant			
Skilled Operator	2.0	8.0			
Operator	10.0	40.0			
Foreman	1.0	4.0			
Lab Tech's etc	3.0	12.0			
TOTAL Operating Jobs	16.0	64.0			
				<u>Annual Cost</u>	
				\$	
Annual Operating Labor Cost				\$28,934,630	
Maintenance Labor Cost				\$47,199,910	
Administration & Support Labor				\$19,033,635	
Property Taxes and Insurance				\$103,736,066	
TOTAL FIXED OPERATING COSTS				\$198,904,241	
VARIABLE OPERATING COSTS					
Maintenance Material Cost					
				\$87,656,976	
<u>Consumables</u>	<u>Consumption</u>		<u>Unit</u>	<u>Initial Fill</u>	
	<u>Initial</u>	<u>/Day</u>	<u>Cost</u>	<u>Cost</u>	
Water/(1000 gallons)	0	12,495	1.67	\$0	\$6,870,522
Chemicals					
MJ & WT Chem (lb)	0	74434	0.27	\$0	\$6,549,255
Carbon (Hg Removal) (lb)	301,013	515	1.63	\$489,217	\$275,185
WDP Sorbent (lb)	Proprietary	Proprietary	Proprietary	\$14,764,440	\$10,522,972
FT Catalyst (lbs)	4,346,525	4347	20.00	\$86,930,497	\$28,556,668
Water Gas Shift Catalyst (ft3)	9,990	5	394.38	\$3,939,872	\$709,177
Methanol (tons)	-	0	300.00	\$0	\$0
AACRP Solvent (gallons)	301,200	83	23.38	\$7,042,056	\$633,785
Hydrotreating catalyst (ft3)	773	0.71	700.00	\$540,835	\$162,187
DSRP catalyst (lbs)	214,720	117.65	11.15	\$2,394,128	\$430,925
Oxidation/Reduction catalyst (lbs)	211,200	115.73	1.56	\$329,472	\$59,307
Calcium Hydroxide (lbs)	5,200	234,926	0.04	\$199	\$2,948,876
Sodium bicarbonate ton)	566	145	151.00	\$85,481	\$7,192,508
Zinc Oxide (ton)	689	0.38	7272.73	\$5,009,455	\$901,702
LTS Catalyst (ton)	14	0.01	22727.27	\$322,727	\$58,091
Subtotal Chemicals			\$121,848,379	\$59,000,638	
Other					
Butane (tons)	0	0	651.34	\$0	\$0
Gases, N2 etc./(/100scf)	0	0	0.00	\$0	\$0
LP Steam (/1000 lbs)	0	0	0.00	\$0	\$0
Subtotal Other			\$0	\$0	
Waste Disposal:					
Spent Mercury Catalyst (lb)	0	515	0.65	\$0	\$110,074
Flyash (ton)	0	0	0.00	\$0	\$0
Slag (ton)	0	2104	25.11	\$0	\$17,354,113
Subtotal Waste Disposal			\$0	\$17,464,187	
By-products & Emissions					
Sulfur (tons)	0	0.0	0.00	\$0	\$0
Supplemental Electricity (MWh generated)	0	1684	-60.00	\$0	-\$33,191,135
Subtotal By-Products			\$0	-\$33,191,135	
TOTAL VARIABLE OPERATING COSTS				\$121,848,379	\$137,801,188
Coal (tons)	0	21,006	68.60	\$0	\$473,372,311

5.6 FISCHER-TROPSCH PRODUCT REQUIRED SELLING PRICE

Table 5-10 shows a summary of the power output, CAPEX, OPEX, and F-T products RSP for the Case 5B Plant. The Case 5B Plant F-T Diesel COP is estimated to be \$131.0/bbl under the loan guarantee finance structure and \$152.0/bbl under the commercial fuels finance structure.

Table 5-10
Case 5B Overall Performance and Economic Summary

Case	Case 5B	
Gasifier	Shell	
Sulfur and CO ₂ Removal	RTI WDP/AACRP	
CAPEX, \$MM		
Total Installed Cost (TIC)	\$3,734	
Total Plant Cost (TPC)	\$5,187	
Total Overnight Cost (TOC)	\$6,530	
OPEX, \$MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC _{fix})	\$198.9	
Variable Operating Cost, less Fuel (OC _{var})	\$190.0	
Coal Fuel (OC _{fuel})	\$526.0	
Import/(Export) Power (OC _{power})	(\$36.9)	
Total OPEX	\$878.0	
Plant Output		
Crude FT Diesel Product, bbl/day	34,994	
Crude FT Naphtha Product, bbl/day	15,006	
Net Power Output, MWe	70.2	
Power Generation		
Gas Turbine Power	101.5	
Steam Turbine Power	448.6	
Syngas Turboexpansion Power	10.0	
Total Gross Power	560.0	
Net Power Generation/(Consumption), MWe	70.2	
Common Factors Used for RSP Calculation		
Capital Charge Factor (CCF)	0.1700	0.218
Capacity Factor (CF)	0.9	
Required Selling Price^{A,B,C}		
RSP FT Diesel, \$/bbl _{FTD}	131.0	152.0
RSP FT Naphtha, \$/bbl _{FTN}	89.9	104.6
Equivalent Crude Oil Price, \$/bbl _{ECO}	113.7	131.9
Equivalent Petroleum Diesel Price, \$/bbl _{EPD}	142.1	164.9
F-T Diesel Required Selling Price Breakdown		
CAPEX	74.27	95.25
Fixed O&M	13.31	13.31
Variable O&M	11.44	11.44
Fuel	31.67	31.67
Power Generation	-2.22	-2.22
CO ₂ T&S	2.52	2.52
Total	131.00	151.97
CO₂ Capture		
CO ₂ Capture Product, tpd	25,280	
Total Carbon Captured (as % Carbon in Fuel)	92.8%	
Costs of CO₂ Captured^{A,B,C} (\$/ton)	13.44	14.49

^A Capacity factor assumed to be 90 percent

^B Values shown are for two financial structures

The first (lower value) is based on the loan guarantee finance structure

The second (higher value) is based on the commercial fuels finance structure

^C Includes CO₂ T&S

The Case 5C Aerojet Gasifier with RTI WDP CTL plant, like the Case 5A reference plant, uses Illinois #6 coal and is designed to generate enough syngas to produce a nominal 50,000 bbl/day of FT liquids product. The Case 5C plant is equipped with gas and steam turbines to generate power to meet the plant's auxiliary power demands. No power import is required, and excess power is exported to the grid as byproduct for additional revenue. The Case 5C plant is designed to capture CO₂ with a carbon capture efficiency of more than 90% of the carbon in the coal.

The Case 5C plant is assumed to operate with an annual on-stream capacity factor of 90% or 7,884 hrs/year at full capacity.

The Case 5C plant has the following characteristics that differentiate it from the reference Case 5A plant:

- The Aerojet DSP feed system replaces the reference case's (Case 5A) lockhopper system used to feed dried coal to the gasifier. RTI and Nexant have evaluated the AR DSP feed system in comparison to the Shell lockhopper feed system in the previous studies on this project for DOE on behalf of AR. The results from these earlier studies have been used to establish the performance and cost of the DSP feed system for this case.
- The Aerojet gasifier replaces the Shell gasifier from Case 5A.
- Syngas leaving the gasifier that has been quenched and filtered enters the HCl guard bed, containing sodium bicarbonate, for HCl removal at 750°F (note that such a guard bed is only required for WDP for this case because of the high level of HCl found in syngas produced from Illinois #6 coal and would not normally be needed for most other coals).
- After the chlorine removal, syngas enters the RTI's WDP for sulfur removal at about 750°F instead of going through a low temperature scrubber as in Case 5A.
- After sulfur removal in the WDP, a portion of the treated syngas then enters the AFWGS process. Instead of the sour shift WGS process used in Case 5A, the Case 5C AFWGS process utilizes sweet shift catalyst, which can operate at a higher temperature than the sour shift catalyst and at about half the cost.
- After the syngas is shifted to a 2:1 H₂/CO ratio and cooled, it enters the AACRP for CO₂ capture. Unlike Rectisol[®], which is a physical solvent, activated methyldiethanolamine is a chemical solvent. As ~99.9% of the sulfur compounds have been removed by the WDP process upstream already, the amine process only has to remove CO₂ and is less complicated than the two-stage Rectisol[®] setup. However, the AACRP unit does also capture ~99% of any residual sulfur left in the syngas following WDP along with the CO₂, resulting in an overall system total sulfur reduction of >99.99% (sub-ppmv total sulfur in the final cleaned syngas).

-
- In Case 5C, the DSRP replaces the Claus process, in Case 5A, for sulfur recovery. In the DSRP, sulfur leaving the WDP process in the form of SO₂ is reduced by a slipstream of shifted syngas, forming elemental sulfur, H₂S and COS. The elemental sulfur is condensed while the remaining H₂S and COS are re-oxidized in the presence of air to form SO₂. The SO₂ is then removed in a lime scrubber, forming gypsum (CaSO₄·2H₂O). This combined approach for sulfur capture results in very low net SO₂ emissions.

Due to the different cold gas efficiency of the Aerojet gasifier, the Aerojet gasifier produces higher amount of syngas for the same coal feed when compared with the Shell gasifier in Case 5A. The different CTL plant cases were designed with same coal feed of 21,006 tpd of Illinois #6. The excess syngas produced due to the higher efficiency of the Aerojet gasifier was utilized to produce FT liquids in excess of the nominal capacity of 50,000 bbl/day. The unreacted syngas after the FT process is directed as fuel gas to gas turbines for power generation. Any excess power is exported to the grid as byproduct for additional revenue.

6.1 CASE 5C PROCESS DESCRIPTION

Case 5C is modeled by replacing the Shell gasifier in Case 5A reference model with the AR compact gasifier. The Rectisol[®] AGR removal process, in Case 5A, is replaced with RTI's WDP technology and AACRP for sulfur and CO₂ removal, respectively. The Claus process, in case 5A, for sulfur recovery is replaced by the DSRP in Case 5C. The system description below follows the overall BFD for the Case 5C CTL plant in Figure 6-1. The accompanying stream flows are shown in Table 6-1. Additional descriptions of the Case 5C plant's various processes are provided below.

Aerojet Dry Solids Pump (DSP) Coal Feed System

Dried coal from the atmospheric storage silo enters the Aerojet DSPs via gravity flow. Three DSPs, each with a nominal capacity of 1,000 tons per day (tpd), are required to service each of the gasification trains. The DSPs increase the pressure of the coal from atmospheric to 700 psia and subsequently discharge the coal continuously to a pressurized feed bin.

Coal is continuously withdrawn from the pressurized feed bin and conveyed by HP CO₂ via a single feed line to each gasifier. To maximize conversion efficiency of fuel to syngas in the gasifier, Aerojet uses its proprietary static splitter system.

Aerojet Rocketdyne Compact Gasifier

For the Case 5C plant, seven Aerojet Rocketdyne (AR) gasifier trains operating at 615 psia are needed to generate the required amount of syngas. Each gasification train includes a single gasifier with a nominal capacity of 3,000 tpd sized for a bituminous (Illinois #6) coal feed. Based on recent pilot plant gasifier testing under DE-FE0023577 with Powder River Basin coal, demonstrating >98% conversion in less than 1/3 the residence time associated with the nominal design gasifier capacity, each of these gasifiers is estimated to provide a minimum of 3,600 tpd capacity. This allows six out of seven gasifiers to gasify 18,000 tpd of dried coal per day out of

the total dried coal feed of 19,647 tpd for Case 5C, providing 92% capacity factor without an installed spare.

Fuel feeds from the pressurized feed bin via the dense phase feed line, conveyed by HP CO₂. To maximize conversion efficiency of fuel to syngas, AR splits the feedstock from a single feed line into many small injection ports via its proprietary static splitter system. The injection ports maximize mixing of coal and oxygen to initiate the gasification reaction.

The AR gasifier is oriented in a vertical, down-firing position. The gasifier reaction is initiated with a torch burner, which is ignited at full gasifier pressure. The ignition torch runs on natural gas and oxygen.

The gasifier injector faceplate and the gasifier liner are water cooled to maintain the metal components at temperatures conducive to long life. The cooling water needs to be clean enough (HP BFW quality) to prevent scale buildup or clogging of internal cooling passages.

The gasifier's raw syngas product is partially quenched from about 2,350°F to around 800°F through the introduction of quench water spray. The quench water enters the gasifier through multiple hydraulic atomizing spray nozzles.

The solids are removed as slag and ash. Liquid slag is solidified in a water bath and removed via a lock hopper system. Ash carried over with the syngas is removed in a candle filter. The ash is also removed via a lock hopper system.

Dry Solids Removal

Contrary to Case 5A, the raw syngas exiting the ceramic particulate filter enters an HCl guard bed for the removal of chlorides. The HCl guard bed uses sodium bicarbonate, which reacts with HCl forming NaCl, carbon dioxide and H₂O, for chloride removal. Chloride removal is achieved in four packed beds, which removes essentially all traces of entrained particles and chlorine. Note that such a guard bed is only required for WDP for this case because of the high level of HCl found in syngas produced from Illinois #6 coal and would not normally be needed for most other coals.

Thus, unlike in Case 5A, the syngas scrubber is eliminated in Case 5C as the syngas leaving the raw syngas coolers at around 750°F enters the HCl guard bed followed by RTI WDP directly without further quench cooling.

Warm Syngas Cleanup Process

The Case 5C warm syngas cleanup process consists of five major system components: RTI WDP, RTI DSRP, AFWGS, LTGC, and the AACRP unit. These are described in greater detail below.

Warm Gas Desulfurization Process

Same as Case 5B.

Direct Sulfur Recovery Process

Same as Case 5B.

Water Gas Shift

Same as Case 5B.

Low Temperature Gas Cooling

Same as Case 5B.

Activated Amine CO₂ Removal Process

Same as Case 5B.

Fischer Tropsch Synthesis and Product Upgrade

FT synthesis and product upgrade for the Case 5C plant is very similar to Case 5A and Case 5B. Due to its higher efficiency compared to the Shell gasifier, the AR gasifier produces a higher amount of syngas production that therefore results in higher FT liquids production in Case 5C.

Figure 6-1
Case 5C Plant – Overall BFD

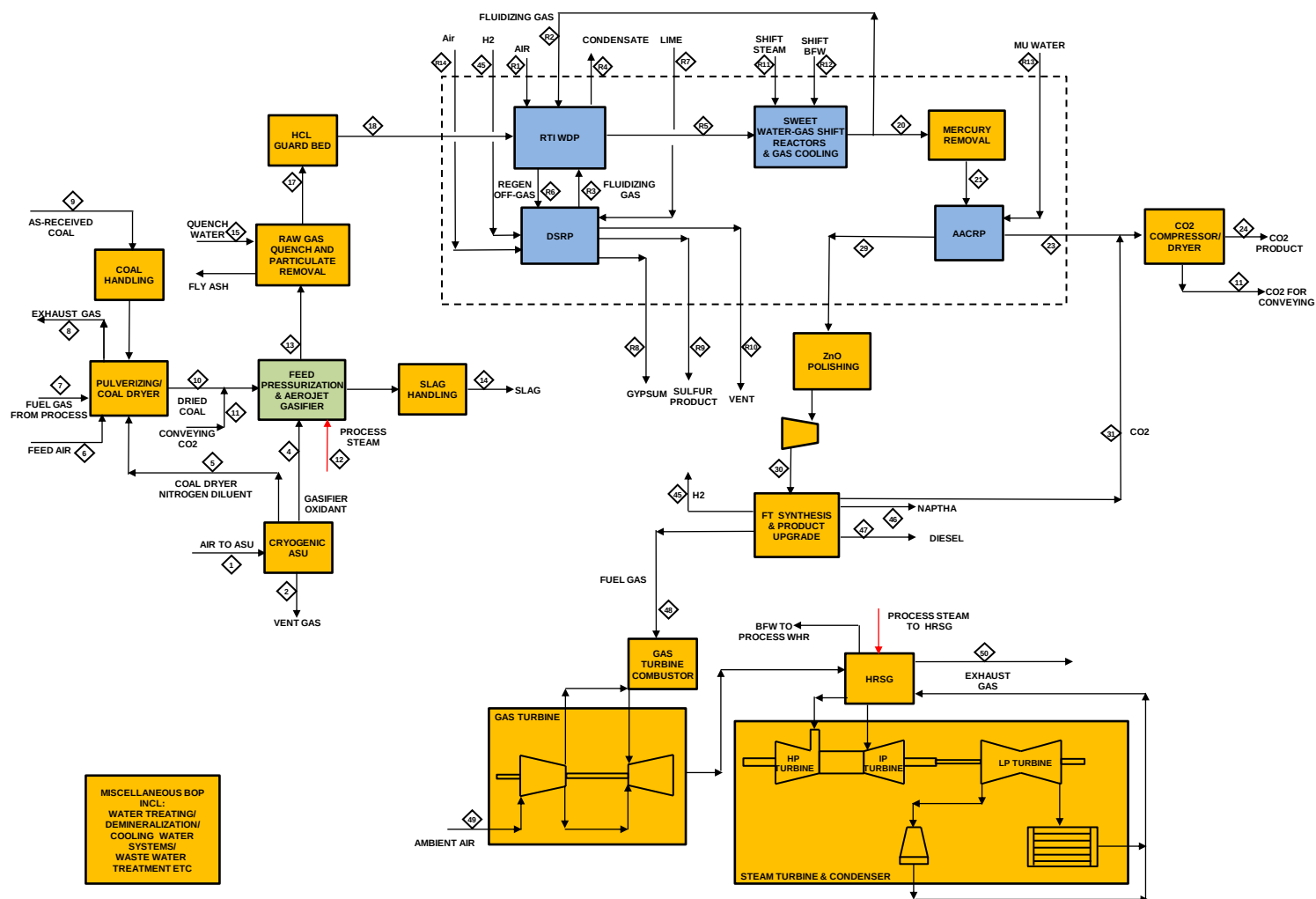


Table 6-1
Case 5C Plant –Overall Stream Table

	1 Air to ASU	2 Vent Gas from ASU	4 Gasifier Oxidant	5 Coal Dryer Nitrogen Diluent	6 Air to Coal Dryer	7 Fuel Gas to Coal Dryer	8 Exhaust Gas from Coal Dryer	9 As-Received Coal	10 Dried Coal	11 Conveying CO2	12 Gasifier Steam	13 Raw Syngas
Mole Frac												
H2	0.0000	0.0000	0.0000	0.0000	0.0000	0.1285	0.0000	0.0000	0.0000	0.0000	0.0000	0.3142
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0179	0.0000	0.0000	0.0000	0.0000	0.0000	0.5858
CO2	0.0003	0.0077	0.0000	0.0000	0.0003	0.2013	0.0105	0.0000	0.0000	1.0000	0.0000	0.0253
CH4	0.0000	0.0000	0.0000	0.0000	0.0000	0.1630	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004
H2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0080
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0007
NH3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005
HCL	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0009
N2	0.7732	0.7077	0.0178	0.9919	0.7732	0.2215	0.9024	0.0000	0.0000	0.0000	0.0000	0.0091
AR	0.0092	0.0225	0.0318	0.0023	0.0092	0.1412	0.0047	0.0000	0.0000	0.0000	0.0000	0.0078
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0099	0.0714	0.0000	0.0004	0.0099	0.0049	0.0711	0.0000	0.0000	0.0000	1.0000	0.0473
O2	0.2074	0.1906	0.9504	0.0054	0.2074	0.0000	0.0113	0.0000	0.0000	0.0000	0.0000	0.0000
SO2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Cl2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0028	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0215	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0205	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0263	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0177	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0121	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0200	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
GYPSUM	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
LIME	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mole Flow lbmol/hr	188,279	7,973	38,681	90,785	10,939	1,480	109,525			3,657	14,995	157,678
Mass Flow lb/hr	5,432,960	226,851	1,244,780	2,547,290	315,663	44,147	3,019,870	1,750,520	1,637,060	160,943	270,137	3,143,250
Temperature F	59	89.6	241	70	59	87	180	59	60	376	650	2600
Pressure psia	15	190	940	18	15	15	15	15	840	785	740	615
Solid Frac	0	0	0	0	0	0	0	1	1	0	0	0
Enthalpy Btu/hr	-2.29E+08	-8.00E+07	3.90E+07	-8.59E+06	-1.32E+07	-6.53E+07	-9.25E+08	-1.58E+09	-8.08E+08	-6.10E+08	-1.50E+09	-2.76E+09
Average MW	28.86	28.45	32.18	28.06	28.86	29.82	27.57			44.01	18.02	19.93

Table 6-1 (cont'd)
Case 5C Plant – Overall Stream Table

	14 Slag	15 Quench Water	17 Syngas to HCl Removal	18 Syngas to Shift	20 Syngas to Hg Removal	21 Syngas to AACRP	23 Total Captured CO2	29 Clean Syngas from AACRP	30 Treated Syngas to FT	45 Hydrogen to DSRP	R1 Air to WDP	R2 Fluidizing Gas to WDP
Mole Frac												
H2	0.0000	0.0000	0.1862	0.1861	0.4892	0.4892	0.0009	0.6523	0.6523	0.9900	0.0000	0.4892
CO	0.0000	0.0000	0.3471	0.3469	0.2352	0.2352	0.0003	0.3136	0.3136	0.0000	0.0000	0.2352
CO2	0.0000	0.0000	0.0150	0.0155	0.2579	0.2579	0.9078	0.0073	0.0073	0.0000	0.0003	0.2579
CH4	0.0000	0.0000	0.0002	0.0002	0.0003	0.0003	0.0000	0.0004	0.0004	0.0000	0.0000	0.0003
H2S	0.0000	0.0000	0.0047	0.0047	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS	0.0000	0.0000	0.0004	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH3	0.0000	0.0000	0.0003	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL	0.0000	0.0000	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N2	0.0000	0.0000	0.0054	0.0054	0.0099	0.0099	0.0000	0.0132	0.0132	0.0100	0.7732	0.0099
AR	0.0000	0.0000	0.0046	0.0046	0.0063	0.0063	0.0000	0.0084	0.0084	0.0000	0.0092	0.0063
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0000	1.0000	0.4355	0.4358	0.0012	0.0012	0.0910	0.0048	0.0048	0.0000	0.0099	0.0012
O2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2074	0.0000
SO2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Cl2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
GYPSUM	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
LIME	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mole Flow lbmol/hr		108,422	266,099	266,244	195,914	195,914	54,469	146,862	146,862	1,273	9,882	3,459
Mass Flow lb/hr	169,673	1,953,250	5,096,500	5,100,190	3,816,220	3,816,220	2,266,070	1,647,750	1,647,750	2,896	285,144	67,376
Temperature F	140	250	750	750	95	95	122	127	200	85	59	95
Pressure psia	615	740	595	585	498	488	21	496	371	390	15	561
Solid Frac	1	0	0	0	0	0	0	0	0	0	0	0
Enthalpy Btu/hr	1987700	-1.31E+10	-1.58E+10	-1.58E+10	-1.08E+10	-1.08E+10	-8.88E+09	-2.40E+09	-2.32E+09	8.56E+04	-1.20E+07	-1.90E+08
Average MW		18.015	19.681	19.685	19.986	19.986	41.682	11.153	11.153	2.276	28.856	19.986

Table 6-1 (cont'd)
Case 5C Plant – Overall Stream Table

	R3 Stripping Gas to WDP	R4 Condensate	R5 Syngas to Shift	R6 WDP Regenerator Off-Gas	R7 Lime	R8 Gypsum	R9 Sulfur	R10 Vent	R11 Shift Steam	R12 Shift BFW	R13 MU Water	R14 Air to DSRP
Mole Frac												
H2	0.0000	0.0000	0.1896	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.3448	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	0.0000	0.0000	0.0190	0.0003	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003
CH4	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH3	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N2	0.9812	0.0000	0.0073	0.8475	0.0000	0.0002	0.0000	0.9808	0.0000	0.0000	0.0000	0.7788
AR	0.0116	0.0000	0.0047	0.0101	0.0000	0.0000	0.0000	0.0116	0.0000	0.0000	0.0000	0.0093
HCN	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0071	1.0000	0.4341	0.0033	0.9737	0.9794	0.0000	0.0076	1.0000	1.0000	1.0000	0.0028
O2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2089
SO2	0.0000	0.0000	0.0000	0.1388	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Cl2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
IC4H10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000
GYPSUM	0.0000	0.0000	0.0000	0.0000	0.0000	0.0191	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
LIME	0.0000	0.0000	0.0000	0.0000	0.0263	0.0010	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mole Flow		71	270,215	9,844	4,344	5,689	575	3,704	11,550	6,589	5,417	619
lbmol/hr	1,228											
Mass Flow lb/hr	34,497	1,273	5,160,010	325,920	84,659	116,316	18,431	104,004	208,081	118,696	97,596	17,896
Temperature F	171	95.14193	679.9108	888.2352	75	215.7153	292.1936	126.81113	550	490	100	269.46
Pressure psia	415	136	563	556	15	15	15	15	615	650	500	589
Solid Frac	0	0	0	0	0	0	0	0	0	0	0	0
Enthalpy Btu/hr	-1.90E+05	-8.72E+06	-1.63E+10	-1.17E+08	-5.68E+08	-7.25E+08	1.29E+06	-1.66E+06	-1.17E+09	-7.58E+08	-6.68E+08	5.55E+05
Average MW	28.08	18.02	19.10	33.11	19.49	20.03	245.14	28.08	18.02	18.02	18.02	28.93

Figure 6-2
Case 5C Plant –FT Process BFD

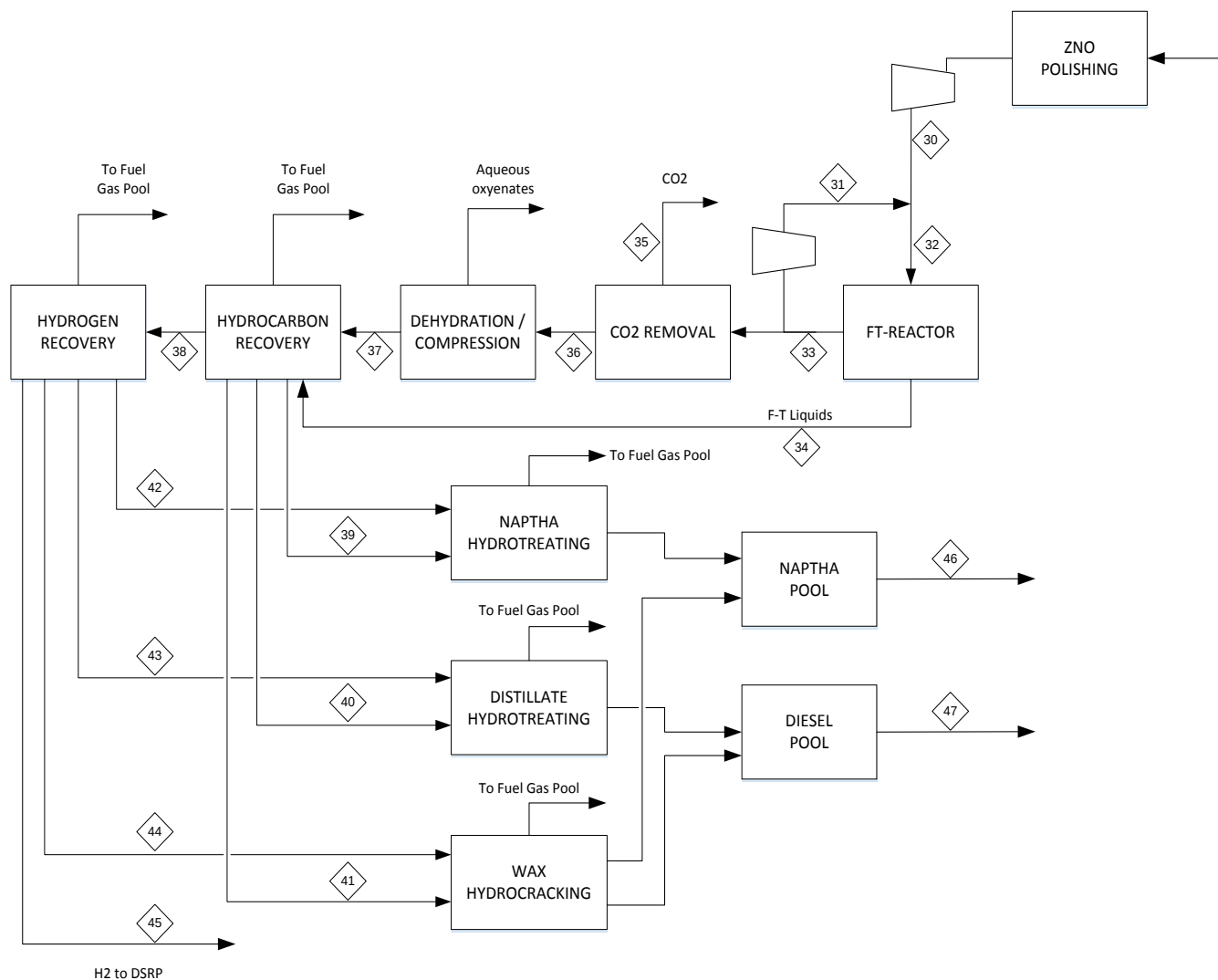


Table 6-2
Case 5C Plant –FT Process Stream Table

	30	31	32	33	34	35	36	37	38	39
Component Mole Fraction										
O2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2	0.01319	0.16129	0.04412	0.16129	0.00368	0.00000	0.17644	0.17708	0.16129	0.00000
CO	0.31364	0.15608	0.28073	0.15608	0.00581	0.00000	0.17075	0.17136	0.15608	0.00000
CO2	0.00731	0.08631	0.02381	0.08631	0.01428	1.00000	0.00047	0.00047	0.08631	0.00000
H2	0.65227	0.31907	0.58267	0.31907	0.00485	0.00000	0.34904	0.35030	0.31907	0.00000
H2O	0.00478	0.00328	0.00446	0.00328	0.06018	0.00000	0.00359	0.00000	0.00328	0.00000
NH3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCN	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H2S	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
AR	0.00841	0.10250	0.02806	0.10250	0.00360	0.00000	0.11213	0.11254	0.10250	0.00000
CH4	0.00040	0.11442	0.02422	0.11442	0.00614	0.00000	0.12517	0.12562	0.11442	0.00000
C2H4	0.00000	0.00195	0.00041	0.00195	0.00033	0.00000	0.00213	0.00214	0.00195	0.00000
C2H6	0.00000	0.01008	0.00210	0.01008	0.00242	0.00000	0.01102	0.01106	0.01008	0.00000
C3H6	0.00000	0.01311	0.00274	0.01311	0.00832	0.00000	0.01434	0.01439	0.01311	0.00000
C3H8	0.00000	0.00592	0.00124	0.00592	0.00428	0.00000	0.00648	0.00650	0.00592	0.00000
IC4H8	0.00000	0.00045	0.00009	0.00045	0.00084	0.00000	0.00050	0.00050	0.00045	0.00000
C4H8	0.00000	0.00907	0.00189	0.00907	0.01698	0.00000	0.00992	0.00996	0.00907	0.00000
IC4H10	0.00000	0.00022	0.00005	0.00022	0.00035	0.00000	0.00024	0.00024	0.00022	0.00000
C4H10	0.00000	0.00397	0.00083	0.00397	0.00868	0.00000	0.00435	0.00436	0.00397	0.00000
C5H10	0.00000	0.00259	0.00054	0.00259	0.01399	0.00000	0.00283	0.00284	0.00259	0.06619
NC5H12	0.00000	0.00441	0.00092	0.00441	0.02869	0.00000	0.00483	0.00484	0.00441	0.12548
IC5H12	0.00000	0.00057	0.00012	0.00057	0.00283	0.00000	0.00062	0.00063	0.00057	0.01393
C6H12	0.00000	0.00114	0.00024	0.00114	0.01798	0.00000	0.00124	0.00125	0.00114	0.06000
NC6H14	0.00000	0.00182	0.00038	0.00182	0.03553	0.00000	0.00199	0.00200	0.00182	0.11373
IC6H14	0.00000	0.00026	0.00005	0.00026	0.00371	0.00000	0.00028	0.00028	0.00026	0.01263
C7H14	0.00000	0.00038	0.00008	0.00038	0.01910	0.00000	0.00042	0.00042	0.00038	0.05434
C7H16	0.00000	0.00068	0.00014	0.00068	0.04078	0.00000	0.00074	0.00074	0.00068	0.11442
C8H16	0.00000	0.00011	0.00002	0.00011	0.01832	0.00000	0.00012	0.00012	0.00011	0.04918
C8H18	0.00000	0.00019	0.00004	0.00019	0.03874	0.00000	0.00021	0.00021	0.00019	0.10357
C9H18	0.00000	0.00003	0.00001	0.00003	0.01690	0.00000	0.00003	0.00003	0.00003	0.04456
C9H20	0.00000	0.00005	0.00001	0.00005	0.03564	0.00000	0.00006	0.00006	0.00005	0.09382
C10H20	0.00000	0.00001	0.00000	0.00001	0.01540	0.00000	0.00001	0.00001	0.00001	0.04038
C10H22	0.00000	0.00001	0.00000	0.00001	0.03244	0.00000	0.00002	0.00002	0.00001	0.08502
C11H22	0.00000	0.00000	0.00000	0.00000	0.00143	0.00000	0.00000	0.00000	0.00000	0.00000

Table 6-2 (cont'd)
Case 5C Reference Plant – FT Process Stream Table

	30	31	32	33	34	35	36	37	38	39
Component Mole										
C11H24	0.00000	0.00001	0.00000	0.00001	0.04195	0.00000	0.00001	0.00001	0.00001	0.00000
C12H24	0.00000	0.00000	0.00000	0.00000	0.00130	0.00000	0.00000	0.00000	0.00000	0.00000
C12H26	0.00000	0.00000	0.00000	0.00000	0.03802	0.00000	0.00000	0.00000	0.00000	0.00000
C13H26	0.00000	0.00000	0.00000	0.00000	0.00118	0.00000	0.00000	0.00000	0.00000	0.00000
C13H28	0.00000	0.00000	0.00000	0.00000	0.03446	0.00000	0.00000	0.00000	0.00000	0.00000
C14H28	0.00000	0.00000	0.00000	0.00000	0.00107	0.00000	0.00000	0.00000	0.00000	0.00000
C14H30	0.00000	0.00000	0.00000	0.00000	0.03122	0.00000	0.00000	0.00000	0.00000	0.00000
C15H30	0.00000	0.00000	0.00000	0.00000	0.00097	0.00000	0.00000	0.00000	0.00000	0.00000
C15H32	0.00000	0.00000	0.00000	0.00000	0.02828	0.00000	0.00000	0.00000	0.00000	0.00000
C16H32	0.00000	0.00000	0.00000	0.00000	0.00087	0.00000	0.00000	0.00000	0.00000	0.00000
C16H34	0.00000	0.00000	0.00000	0.00000	0.02562	0.00000	0.00000	0.00000	0.00000	0.00000
C17H34	0.00000	0.00000	0.00000	0.00000	0.00079	0.00000	0.00000	0.00000	0.00000	0.00000
C17H36	0.00000	0.00000	0.00000	0.00000	0.02321	0.00000	0.00000	0.00000	0.00000	0.00000
C18H36	0.00000	0.00000	0.00000	0.00000	0.00072	0.00000	0.00000	0.00000	0.00000	0.00000
C18H38	0.00000	0.00000	0.00000	0.00000	0.02102	0.00000	0.00000	0.00000	0.00000	0.00000
C19H38	0.00000	0.00000	0.00000	0.00000	0.00065	0.00000	0.00000	0.00000	0.00000	0.00000
C19H40	0.00000	0.00000	0.00000	0.00000	0.01904	0.00000	0.00000	0.00000	0.00000	0.00000
C20H40	0.00000	0.00000	0.00000	0.00000	0.00059	0.00000	0.00000	0.00000	0.00000	0.00000
C20H42	0.00000	0.00000	0.00000	0.00000	0.01724	0.00000	0.00000	0.00000	0.00000	0.00000
C21OP	0.00000	0.00000	0.00000	0.00000	0.01653	0.00000	0.00000	0.00000	0.00000	0.00000
C22OP	0.00000	0.00000	0.00000	0.00000	0.01532	0.00000	0.00000	0.00000	0.00000	0.00000
C23OP	0.00000	0.00000	0.00000	0.00000	0.01420	0.00000	0.00000	0.00000	0.00000	0.00000
C24OP	0.00000	0.00000	0.00000	0.00000	0.01316	0.00000	0.00000	0.00000	0.00000	0.00000
C25OP	0.00000	0.00000	0.00000	0.00000	0.01220	0.00000	0.00000	0.00000	0.00000	0.00000
C26OP	0.00000	0.00000	0.00000	0.00000	0.01131	0.00000	0.00000	0.00000	0.00000	0.00000
C27OP	0.00000	0.00000	0.00000	0.00000	0.01048	0.00000	0.00000	0.00000	0.00000	0.00000
C28OP	0.00000	0.00000	0.00000	0.00000	0.00971	0.00000	0.00000	0.00000	0.00000	0.00000
C29OP	0.00000	0.00000	0.00000	0.00000	0.00900	0.00000	0.00000	0.00000	0.00000	0.00000
C30WAX	0.00000	0.00000	0.00000	0.00000	0.11712	0.00000	0.00000	0.00000	0.00000	0.00000
OXVAP	0.00000	0.00000	0.00000	0.00000	0.00251	0.00000	0.00000	0.00000	0.00000	0.00000
OXHC	0.00000	0.00002	0.00000	0.00002	0.01803	0.00000	0.00002	0.00002	0.00002	0.02274
OXH2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
GASOLINE	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
DIESEL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Mole Flow (LBMOL/HR)	146,862	38,776	185,638	50,701	2,729	1,024	10,901	10,861	38,776	1,043
Mass Flow (LB/HR)	1,647,740	862,469	2,510,210	1,127,710	576,327	45,072	220,164	219,459	862,469	107,086
Temperature (F)	200	149	187	100	286	100	100	100	100	257
Pressure (PSIA)	371	378	371	284	75	275	275	430	284	50

Table 6-2 (cont'd)
Case 5C Reference Plant – FT Process Stream Table

	40	41	42	43	44	45	46	47	51	52
Component Mole Fraction										
O2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.16098
CO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.01050
CO2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.14507
H2	0.00000	0.00000	1.00000	1.00000	1.00000	1.00000	0.00000	0.00000	0.00000	0.46472
H2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.00000	0.00153
NH3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCN	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H2S	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
AR	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.10106
CH4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.10991
C2H4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00115
C2H6	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00386
C3H6	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00084
C3H8	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00030
IC4H8	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H8	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00005
IC4H10	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H10	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00002
C5H10	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NC5H12	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.07035	0.00000	0.00000	0.00000
IC5H12	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.03069	0.00000	0.00000	0.00000
C6H12	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NC6H14	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.15203	0.00000	0.00000	0.00000
IC6H14	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.05373	0.00000	0.00000	0.00000
C7H14	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H16	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.43985	0.00000	0.00000	0.00000
C8H16	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H18	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.06050	0.00000	0.00000	0.00000
C9H18	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H20	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H20	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H22	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C11H22	0.00509	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000

Table 6-2 (cont'd)
Case 5C Reference Plant – FT Process Stream Table

	40	41	42	43	44	45	46	47	51	52
Component Mole Fraction										
C11H24	0.14928	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C12H24	0.00461	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C12H26	0.13522	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C13H26	0.00418	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C13H28	0.12254	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C14H28	0.00379	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C14H30	0.11100	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C15H30	0.00343	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C15H32	0.10056	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C16H32	0.00311	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C16H34	0.09108	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C17H34	0.00282	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C17H36	0.08251	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C18H36	0.00255	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C18H38	0.07473	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C19H38	0.00231	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C19H40	0.06769	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H40	0.00000	0.00238	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H42	0.00000	0.06985	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C21OP	0.00000	0.06696	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C22OP	0.00000	0.06207	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C23OP	0.00000	0.05753	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C24OP	0.00000	0.05333	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C25OP	0.00000	0.04942	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C26OP	0.00000	0.04580	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C27OP	0.00000	0.04245	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C28OP	0.00000	0.03934	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C29OP	0.00000	0.03646	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C30WAX	0.00000	0.47439	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
OXVAP	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
OXHC	0.03348	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
OXH2O	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
GASOLINE	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.19286	0.00000	0.00000	0.00000
DIESEL	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.00000	0.00000	0.00000
Mole Flow (LBMOL/HR)	768	674	518	41	1,314	2,545	1,538	1,891	5,433	11,883
Mass Flow (LB/HR)	153,530	315,015	1,044	84	2,648	5,130	154,946	401,676	97,877	215,722
Temperature (F)	257	257	100	100	602	85	100	100	650	85
Pressure (PSIA)	50	50	50	50	50	385	50	50	615	385

6.2 SPARING PHILOSOPHY

The sparing philosophy for the Aerojet and RTI technologies for the Case 5C CTL plant is provided below. The design has:

- Seven trains of gasification, including dry feed system, Aerojet gasifier, cyclone and candle filter (7 x 14%)
- Three WDP trains (3 x 33%)
- Three AACRP trains (3 x 33%)
- Two DSRP trains (2 x 50%)
- Two AFWGS trains (2 x 50%)

6.3 PERFORMANCE RESULTS

As discussed in the previous sections, due to the higher efficiency of the AR gasifier, Case 5C plant consumes 21,006 tpd of Illinois #6 coal, which is similar to that consumed in Case 5A and Case 5B, at the Midwestern site, but produces 51,601 bbl/day of FT liquid product, containing 70% FT diesel and 30% FT Naphtha. Overall performance for the Case 5C plant is summarized in Table 6-3, which includes auxiliary power requirements. Loads that are unique to the RTI WDP system are shown in bold and italicized.

Table 6-3
Case 5C Plant Performance Summary

Power Cycle Power Generation	Electrical Load, kWe
Gas Turbine Power	104,927
Steam Turbine Power	414,159
Syngas Turboexpansion Power	11,361
Total	530,447
CTL Auxiliary Load Description	Electrical Load, kWe
Coal Handling	1,760
Coal Milling	8,160
Slag Handling	2,080
ASU Auxiliaries	3,570
ASU Main Air Compressor	210,142
Oxygen Compressor	34,707
Quench Water Pump	2,812
RTI WDP	20,792
AACRP	38,760
DSRP	3,083
CO ₂ Compressor Auxiliary	71,312
FT Power Requirement	27,696
Miscellaneous BOP	13,402
Circulating Water Pumps	14,361
Ground Water Pumps	800
Cooling Tower Fans	7,519
Gas Turbine Auxiliaries	4,000
Steam Turbine Auxiliaries	400
Transformer Losses	3,129
Condensate Pumps	200
Boiler Feed Water Pumps	3,371
Total	472,056
NET PLANT POWER, kWe	58,392

Table 6-4 shows the carbon balance for the Case 5C plant. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not part of the carbon capture equation, but is not neglected in the balance since the model accounts for the air components throughout. Carbon leaves the CTL plant as unburned carbon in the sour water and LTGC condensate, slag, CO₂ in the dryer exhaust gas, ASU vent gas, FT products, sulfur products, and both the AACRP CO₂ capture products. Additionally, carbon also leaves the power

cycle as CO₂ in the stack. Carbon in the F-T products is considered to be product and not emissions. The carbon capture efficiency is defined as the amount of carbon in the product streams, which include the FT and the dried and compressed CO₂ products from the Rectisol[®] AGR, relative to the amount of carbon in the coal feedstock less carbon contained in the slag. For Case 5C, the carbon capture efficiency is 92.6%, compared to 92.5% in Case 5A and 92.8% in case 5B.

Table 6-4
Case 5C Plant – Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Coal Feed	1,115,957	
ASU Air	739	
Carbon in during HCl Removal (as NaHCO ₃)	1,736	
Coal Dryer Air	39	
Carbon in Air to WDP	36	
Carbon in Air to DSRP	4	
Carbon in Sour water		3,426
Carbon in LTGC condensate		110
ASU Vent		739
Exhaust Gas to Stack		81,169
CO ₂ (post FT)		12,300
Carbon in Slag		-
Sulfur Product		-
Gypsum		38
DSRP Vent		-
FT Naphtha		130,012
FT Diesel		340,656
AACRP CO ₂ Product		550,061
Total	1,118,511	1,118,511

Table 6-5 shows the sulfur balance for the Case 5C plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the elemental sulfur recovered in the DSRP, sulfur recovered in the form of gypsum leaving the lime scrubber, sulfur emitted in the coal dryer exhaust gas, sulfur emitted in the power cycle exhaust gas, sulfur emitted in the FT liquids product, sulfur emitted in the DSRP vent gas and sulfur that is sequestered with the CO₂ product. Sulfur in the slag is considered negligible. There is no net sulfur emission from the Case 5B CTL plant.

Table 6-5
Case 5C Plant – Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Coal Feed	43,855	
Coal Dryer Exhaust Gas		-
GT Exhaust Gas		-
CO ₂ (post FT)		-
Sulfur Product		34,324
Gypsum		9,492
DSRP Vent		-
FT Naphtha		-
FT Diesel		-
AACRP CO ₂ Product		39
Total	43,855	43,855

Table 6-6 shows the overall water balance for the Case 5B plant. Raw water withdrawal is the amount of raw water consumed by the plant. Some water is internally recycled and consumed by the slag handling process, syngas quench process and as makeup to the cooling tower.

Table 6-6: Case 5C plant – Overall Water Balance

Water Usage by Area	Water Demand		Internal Recycle		Raw water Withdrawal		Process Water Discharge		Raw Water Consumption	
	m ³ /min	gpm	m ³ /min	gpm	m ³ /min	gpm	m ³ /min	gpm	m ³ /min	gpm
Slag Handling	1.7	458	1.7	458	-	-	-	-	-	-
Quench/Wash	14.8	3,899	10.4	2,736	4.4	1,162	-	-	4.4	1,162
AGR Makeup Water	0.7	195	-	-	0.7	195	-	-	0.7	195
SWS Blowdown	-	-	-	-	-	-	0.1	30	(0.1)	(30)
FT-Area	0.1	39	-	-	0.1	39	-	-	0.1	39
Condenser Makeup	5.8	1,519		-	5.8	1,519		-	5.8	1,519
Gasifier Steam	2.0	539	-	-	2.0	539	-	-	2.0	539
Shift Steam	3.2	848	-	-	3.2	848	-	-	3.2	848
BFW Makeup	0.5	132	-	-	0.5	132	-	-	0.5	132
Cooling Tower	56.6	14,948	-	-	56.6	14,948	14.1	3,730	42.5	11,218
BFW Blowdown	-	-	0.5	132	(0.5)	(132)	-	-	(0.5)	(132)
SWS Blowdown	-	-	1.1	302	(1.1)	(302)	-	-	(1.1)	(302)
TOTAL	79.8	21,057		3,628	66.0	17,430		3,760	51.8	13,669

6.4 CAPITAL COST

Table 6-7 shows the cost breakdown of the Case 5C plant, consistent with the Code of Accounts format as expressed in the CTL Study. The accounts/subaccounts of interest for this study are:

- 2.3 Dry Coal Injection System,
- 4 Gasifier & Accessories,
- 5A Gas Cleanup & Piping and
- 5AA F-T Synthesis and Product Upgrade
- 5B CO₂ Removal and Compression

These are shown with more detail to include the various subaccounts and to provide more clarity as to wherein the major cost differences lie among the cases.

Table 6-8 shows the calculation and addition of owner's costs to determine the TOC, which is used to calculate the cost of production for FT products.

The estimated TOC of the Case 5C plant in 2011 dollars is \$5,797 MM.

Table 6-7
Case 5C Plant – Total Plant Cost Summary

Case 5c: Aerojet Gasifier w/ RTI WDP CFTL Plant			Coal Feed, lb/hr 1,750,518				Plant Size		51,602 barrels per day of FT-liquids			
Illinois #6			Coal HHV, Btu/lb 13,126 (dry basis)				Cost Base		Jun 2011 (\$x1000)			
Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost \$	Eng'g CM H.O & Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect				Process	Project	\$	
1	COAL & SORBENT HANDLING	\$38,319	\$6,719	\$29,308	\$0	\$0	\$74,346	\$6,589	\$0	\$16,187	\$97,122	
2	COAL & SORBENT PREP & FEED											
2.3	Dry Coal Injection System	\$188,734	\$0	\$123,925	\$0	\$0	\$312,659	\$26,040	\$62,532	\$80,247	\$481,477	
2.x	Other Coal & Sorbent Prep & Feed Systems	\$130,490	\$22,642	\$35,141	\$0	\$0	\$188,273	\$15,985	\$0	\$40,851	\$245,109	
	SUBTOTAL 2.	\$319,224	\$22,642	\$159,066	\$0	\$0	\$500,932	\$42,025	\$62,532	\$121,098	\$726,586	
3	FEEDWATER & MISC BOP SYSTEMS	\$12,911	\$8,267	\$13,963	\$0	\$0	\$35,141	\$3,256	\$0	\$9,368	\$47,765	
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries	\$132,286	\$66,143	\$199,735	\$0	\$0	\$398,164	\$35,042	\$99,541	\$81,964	\$614,710	
4.3	ASU/Oxidant Compression	\$508,999	\$0 w/equip.	\$0	\$0	\$0	\$508,999	\$47,561	\$0	\$55,656	\$612,216	
4.4	LT Heat Recovery & FG Saturation	\$90,796	\$0	\$28,824	\$0	\$0	\$119,620	\$11,302	\$0	\$26,184	\$157,106	
4.6	Flare Stack System	\$0	\$3,563	\$1,440	\$0	\$0	\$5,003	\$467	\$0	\$1,094	\$6,564	
4.9	Gasification Foundations	\$0	\$43,142	\$25,729	\$0	\$0	\$68,871	\$6,348	\$0	\$18,805	\$94,024	
	SUBTOTAL 4.	\$732,081	\$112,848	\$255,727	\$0	\$0	\$1,100,656	\$100,720	\$99,541	\$183,703	\$1,484,620	
5A	GAS CLEANUP & PIPING											
5A.1a	RTI WDP for Sulfur Removal	\$103,934	\$0	\$124,721	\$0	\$0	\$228,656	\$21,575	\$45,731	\$59,192	\$355,154	
5A.1b	AA CO2 Removal Process	\$73,715	\$0	\$59,901	\$0	\$0	\$133,617	\$12,607	\$26,723	\$34,589	\$207,536	
5A.2	Elemental Sulfur Plant	\$23,585	\$0	\$25,291	\$0	\$0	\$48,876	\$4,642	\$9,775	\$12,659	\$75,953	
5A.3	Mercury Removal	\$2,985	\$0	\$2,256	\$0	\$0	\$5,241	\$494	\$262	\$1,199	\$7,196	
5A.4	Shift Reactors	\$4,794	\$0	\$1,918	\$0	\$0	\$6,712	\$627	\$0	\$1,468	\$8,806	
5A.5	Particulate Removal	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	
5A.6	Blowback Gas Systems	\$3,633	\$611	\$342	\$0	\$0	\$4,586	\$423	\$0	\$1,002	\$6,011	
5A.7	Fuel Gas Piping	\$0	\$1,817	\$1,189	\$0	\$0	\$3,006	\$262	\$0	\$654	\$3,923	
5A.9	HGCU Foundations	\$0	\$1,965	\$1,324	\$0	\$0	\$3,289	\$304	\$0	\$1,078	\$4,671	
5A.10	Syngas Conditioning	\$26,998	\$19,062	\$22,873	\$0	\$0	\$68,934	\$6,549	\$0	\$15,096	\$90,579	
5A.1c	HCl Guard Bed	\$5,886	\$2,633	\$3,408	\$0	\$0	\$11,928	\$1,133	\$0	\$2,612	\$15,673	
	SUBTOTAL 5.	\$245,531	\$26,089	\$243,224	\$0	\$0	\$514,844	\$48,616	\$82,491	\$129,550	\$775,502	
5AA	F-T SYNTHESIS AND PRODUCT UPGRADE											
5AA.1	FT Synthesis*	\$145,056	\$0	\$0	\$0	\$0	\$145,056	\$13,925	\$39,165	\$49,537	\$247,683	
5AA.2	Amine CO2 Absorption	\$6,703	w/equip.	w/equip.	\$0	\$0	\$6,703	\$632	\$1,341	\$1,735	\$10,411	
5AA.4	Compression	\$8,312	\$0	\$0	\$0	\$0	\$8,312	\$798	\$2,244	\$2,839	\$14,193	
5AA.5	Hydrocarbon Recovery	\$35,820	\$0	\$0	\$0	\$0	\$35,820	\$3,439	\$9,671	\$12,232	\$61,163	
5AA.6	Hydrogen Recovery	\$11,368	\$0	\$0	\$0	\$0	\$11,368	\$1,091	\$3,069	\$3,882	\$19,410	
5AA.8	Naphtha Hydrotreater	\$13,521	\$0	\$0	\$0	\$0	\$13,521	\$1,298	\$3,651	\$4,618	\$23,088	
5AA.9	Diesel Hydrotreater	\$33,905	\$0	\$0	\$0	\$0	\$33,905	\$3,256	\$9,154	\$11,579	\$57,895	
5AA.10	Wax Hydrocracker	\$83,253	\$0	\$0	\$0	\$0	\$83,253	\$7,993	\$22,479	\$28,431	\$142,156	
5AA.11	Raw Fuel Gas Compressor, HP	\$23,221	w/equip.	w/equip.	\$0	\$0	\$23,221	\$2,166	\$0	\$5,078	\$30,465	
5AA.12	Fuel Gas to GT Compressor, HP	\$6,120	w/equip.	w/equip.	\$0	\$0	\$6,120	\$571	\$0	\$1,338	\$8,029	
	SUBTOTAL 5AA.	\$367,279	\$0	\$0	\$0	\$0	\$367,279	\$35,168	\$90,775	\$121,269	\$614,491	
5B	CO2 Compression & Drying	\$82,528	\$0	\$31,112	\$0	\$0	\$113,640	\$10,597	\$0	\$24,848	\$149,085	
	SUBTOTAL 5B.	\$82,528	\$0	\$31,112	\$0	\$0	\$113,640	\$10,597	\$0	\$24,848	\$149,085	
6	COMBUSTION TURBINE/ACCESSORIES	\$50,295	\$418	\$4,047	\$0	\$0	\$54,761	\$11,854	\$12,637	\$15,760	\$95,012	
7	HRSG, DUCTING & STACK	\$17,155	\$1,502	\$4,878	\$0	\$0	\$23,536	\$2,170	\$0	\$2,994	\$28,699	
8	STEAM TURBINE GENERATOR	\$64,930	\$1,489	\$18,919	\$0	\$0	\$85,337	\$7,432	\$0	\$12,388	\$105,158	
9	COOLING WATER SYSTEM	\$15,766	\$20,993	\$18,249	\$0	\$0	\$55,008	\$4,936	\$0	\$12,710	\$72,654	
10	ASH/SPENT SORBENT HANDLING SYS	\$50,895	\$3,552	\$25,045	\$0	\$0	\$79,493	\$7,430	\$0	\$9,426	\$96,349	
11	ACCESSORY ELECTRIC PLANT	\$40,202	\$23,597	\$41,167	\$0	\$0	\$104,967	\$8,897	\$0	\$22,623	\$136,487	
12	INSTRUMENTATION & CONTROL	\$15,457	\$3,133	\$10,163	\$0	\$0	\$28,753	\$2,559	\$1,437	\$5,491	\$38,240	
13	IMPROVEMENTS TO SITE	\$6,190	\$8,365	\$24,054	\$0	\$0	\$38,609	\$3,822	\$0	\$12,729	\$55,161	
14	BUILDINGS & STRUCTURES	\$0	\$25,102	\$17,544	\$0	\$0	\$42,646	\$3,790	\$0	\$7,415	\$53,851	
	CALCULATED TOTAL COST	\$2,058,763	\$264,717	\$896,467	\$0	\$0	\$3,219,947	\$299,861	\$349,413	\$707,560	\$4,576,781	

Table 6-8
Case 5C Plant – Total Overnight Cost Breakdown

Owner's Costs	\$ x \$1,000
Preproduction Costs	
6 months All Labor	\$44,115
1 Month Maintenance Materials	\$7,162
1 Month Non-Fuel Consumables	\$6,161
1 Month Waste Disposal	\$1,617
25% of 1 Months Fuel Cost at 100% CF	\$10,958
2% of TPC	\$91,536
Total	\$161,547
Inventory Capital	
60 day supply of fuel at 100% CF	\$86,461
60-day supply of non-fuel consumables at 100% CF	\$12,153
0.5% of TPC (spare parts)	\$22,884
Total	\$121,497
Initial Cost for Catalyst and Chemicals	\$126,248
Land	\$900
Other Owner's Cost	\$683,431
Financing Costs	\$123,018
Total Owner's Costs	\$1,214,697
Total Overnight Costs (TOC)	\$5,797,064
TASC Multiplier	1.147
Total As-Spent Cost (TASC)	\$6,649,233

6.5 OPERATING COSTS

Table 6-9 shows the operating cost breakdown for the Case 5C Plant.

Table 6-9
Case 5C Plant – Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES					
Case:	Aerojet Gasifier w/ RTI WDP CFTL Plant				
Plant Size (bbl/day FT Liquids)	51,602				
Primary/Secondary Fuel:	Illinois #6 Coal				
Design/Construction	5 years		Book Life (yrs):	20	
TPC (Plant Cost) Year	June 2011		TPI Year:	2016	
Capacity Factor (%)	90		CO2 Captured (TPD)	24,729	
OPERATING & MAINTENANCE LABOR					
Operating Labor					
Operating Labor Rate (base):	\$39.70 \$/hr				
Operating Labor Burden:	30.00 % of base				
Labor Overhead Charge	25.00 % of labor				
Operating Labor Requirements per Shift	units/mod		Total Plant		
Skilled Operator	2.0		8.0		
Operator	10.0		40.0		
Foreman	1.0		4.0		
Lab Tech's etc	3.0		12.0		
TOTAL Operating Jobs	16.0		64.0		
			Annual Cost		
			\$		
Annual Operating Labor Cost			\$28,934,630		
Maintenance Labor Cost			\$41,648,711		
Administration & Support Labor			\$17,645,835		
Property Taxes and Insurance			\$91,535,629		
TOTAL FIXED OPERATING COSTS			\$179,764,805		
VARIABLE OPERATING COSTS					
Maintenance Material Cost			\$77,347,606		
Consumables	Consumption		Unit	Initial Fill	
	Initial	/Day	Cost	Cost	
Water/(1000 gallons)	0	12,549	1.67	\$0	\$6,900,215
Chemicals					
MU & WT Chem (lb)	0	74756	0.27	\$0	\$6,577,559
Carbon (Hg Removal) (lb)	301,013	515	1.63	\$489,217	\$275,185
WDP Sorbent (lb)	Proprietary	Proprietary	Proprietary	\$14,727,960	\$10,497,065
FT Catalyst (lbs)	4,485,787	4486	20.00	\$89,715,750	\$29,471,624
Water Gas Shift Catalyst (ft3)	9,872	5	394.38	\$3,893,289	\$700,792
Methanol (tons)	0	0	300.00	\$0	\$0
AACRP Solvent (gallons)	387,600	106	23.38	\$9,062,088	\$815,588
Hydrotreating catalyst (ft3)	773	0.71	700.00	\$540,835	\$162,187
DSRP catalyst (lbs)	214720	117.65	11.15	\$2,394,128	\$430,925
Oxidation/Reduction catalyst (lbs)	211200	115.73	1.56	\$329,472	\$59,307
Calcium Hydroxide (lbs)	5200	203184.00	0.04	\$199	\$2,550,439
Sodium bicarbonate ton)	566	145.00	151.00	\$85,481	\$7,192,508
Zinc Oxide (ton)	689	0	7272.73	\$5,009,455	\$901,702
Subtotal Chemicals				\$126,247,874	\$59,634,881
Other					
Butane (tons)	0	0.0	651.34	\$0	\$0
Gases, N2 etc./(100scf)	0	0	0.00	\$0	\$0
LP Steam (/1000 lbs)	0	0	0.00	\$0	\$0
Subtotal Other				\$0	\$0
Waste Disposal:					
Spent Mercury Catalyst (lb)	0	515	0.65	\$0	\$110,074
Flyash (ton)	0	0	0.00	\$0	\$0
Slag (ton)	0	2104	25.11	\$0	\$17,354,113
Subtotal Waste Disposal				\$0	\$17,464,187
By-products & Emissions					
Sulfur (tons)	0	0.0	0.00	\$0	\$0
Supplemental Electricity (MWh generated)	0	1401	-60.00	\$0	-\$27,621,531
Subtotal By-Products				\$0	-\$27,621,531
TOTAL VARIABLE OPERATING COSTS				\$126,247,874	\$133,725,357
Coal (tons)	0	21,006	68.60	\$0	\$473,372,311

6.6 FISCHER-TROPSCH PRODUCT REQUIRED SELLING PRICE

Table 6-10 shows a summary of the power output, CAPEX, OPEX, and FT products RSP for the Case 5C Plant. The Case 5C Plant FT Diesel RSP is estimated to be \$117.3/bbl under the loan guarantee finance structure and \$135.6/bbl under the commercial fuels finance structure.

Table 6-10
Case 5C Plant – Overall Performance and Economic Summary

Case	Case 5C	
Gasifier	Aerojet	
Sulfur and CO ₂ Removal	RTI WDP/AACRP	
CAPEX, \$MM		
Total Installed Cost (TIC)	\$3,220	
Total Plant Cost (TPC)	\$4,577	
Total Overnight Cost (TOC)	\$5,797	
OPEX, \$MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC _{fix})	\$179.8	
Variable Operating Cost, less Fuel (OC _{var})	\$179.3	
Coal Fuel (OC _{fuel})	\$526.0	
Import/(Export) Power (OC _{Power})	(\$30.7)	
Total OPEX	\$854.3	
Plant Output		
Crude FT Diesel Product, bbl/day	36,115	
Crude FT Naphtha Product, bbl/day	15,486	
Net Power Output, MWe	58.4	
Power Generation		
Gas Turbine Power	104.9	
Steam Turbine Power	414.2	
Syngas Turboexpansion Power	11.4	
Total Gross Power	530.4	
Net Power Generation/(Consumption), MWe	58.4	
Common Factors Used for RSP Calculation		
Capital Charge Factor (CCF)	0.1700 0.218	
Capacity Factor (CF)	0.9	
Required Selling Price^{A,B,C}		
RSP FT Diesel, \$/bbl _{FTD}	117.3	135.6
RSP FT Naphtha, \$/bbl _{FTN}	80.4	93.1
Equivalent Crude Oil Price, \$/bbl _{ECO}	101.8	117.7
Equivalent Petroleum Diesel Price, \$/bbl _{EPD}	127.3	147.1
F-T Diesel Required Selling Price Breakdown		
CAPEX	63.89	81.93
Fixed O&M	11.65	11.65
Variable O&M	10.46	10.46
Fuel	30.69	30.69
Power Generation	-1.79	-1.79
CO ₂ T&S	2.39	2.39
Total	117.29	135.33
CO₂ Capture		
CO ₂ Capture Product, tpd	24,729	
Total Carbon Captured (as % Carbon in Fuel)	92.6%	
Costs of CO₂ Captured^{A,B,C} (\$/ton)	13.47	15.04

^A Capacity factor assumed to be 90 percent

^B Values shown are for two financial structures

The first (lower value) is based on the loan guarantee finance structure

The second (higher value) is based on the commercial fuels finance structure

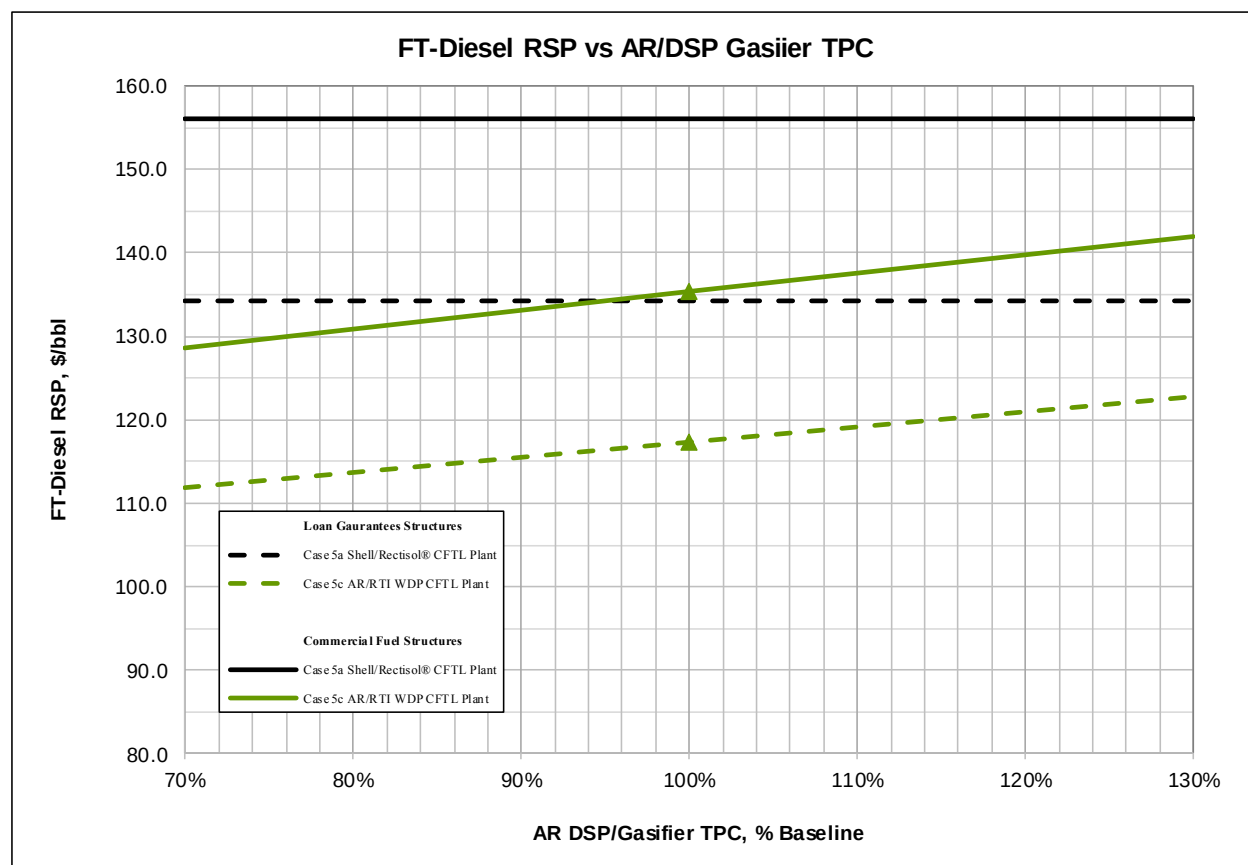
^C Includes CO₂ T&S

Sensitivity analysis was carried out to determine the effects of various parameters of the AR gasifier and RTI WDP systems on the overall FT diesel and FT-naphtha product RSP. Due to the fixed 70:30 split of FT-naphtha and FT-Diesel produced in this study, the trends for the FT-naphtha and FT-Diesel are expected to be similar, so the results are reported for FT-Diesel only. The parameters investigated include: AR and RTI systems capital cost, CTL plant capacity factor, feedstock (coal) cost, electric selling price, CO₂ sales price, cost of CO₂ emissions and capital charge factor.

7.1 AR DSP/GASIFIER SYSTEM COST

Figure 7-1 shows how the FT-Diesel RSPs for the AR gasifier case (5C) change as the AR DSP and gasifier TPC varies from -30% to +30%. Also shown in figure are the reference Case 5A FT-Diesel RSPs at \$156.0/bbl and \$134.3/bbl for the commercial fuels and loan guarantee finance structures respectively.

Figure 7-1
Sensitivity Analysis – RSP vs AR DSP/Gasifier TPC



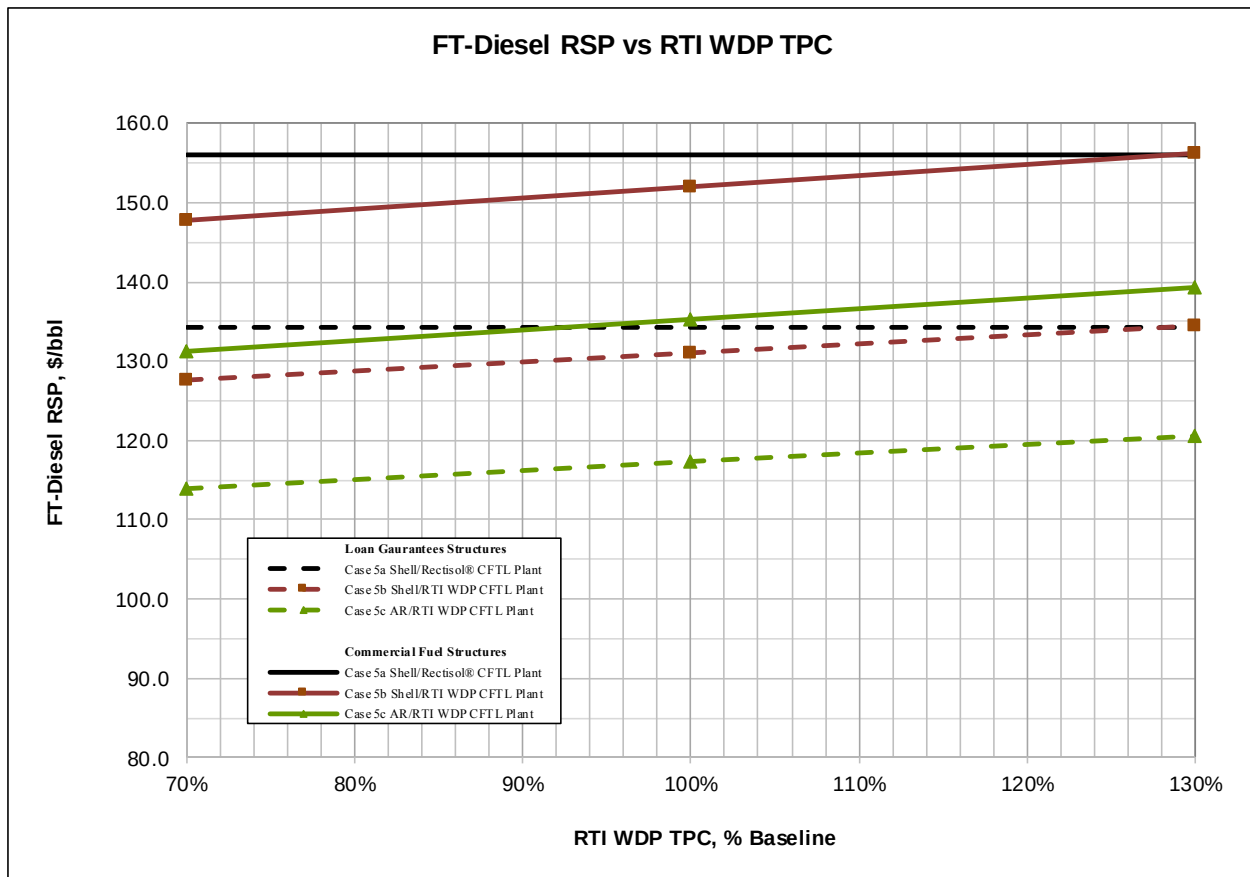
From Figure 7-1, it can be seen that at the high end of the AR DSP and gasifier TPC (+30%), the FT-Diesel RSPs for Cases 5C are still less than the FT-Diesel RSP for the Case 5A reference plant.

Under the commercial fuels financing structure, every 4.5% increase in DSP and gasifier TPC, equivalent to about \$50 MM, increases the FT-Diesel RSP by \$1/bbl. Under the loan guarantee financing structure, every 5.5% increase in DSP and gasifier TPC, equivalent to about \$61 MM, increases the FT-Diesel RSP by \$1/bbl.

7.2 RTI WDP SYSTEM COST

Figure 7-2 shows how FT-Diesel RSPs for the RTI WDP plant cases (Cases 5B and 5C) change as the RTI WDP system TPC varies from -30% to +30%. The RTI WDP system TPC includes the costs for the RTI WDP process, DSRP, AFWGS, AACRP and the HCl guard bed process. For reference purposes, the Case 5A reference plant FT-Diesel RSPs at \$156.0/bbl for the commercial fuels and \$134.3/bbl for the loan guarantee finance structures are shown in Figure 7-2 as well.

Figure 7-2
Sensitivity Analysis – RSP vs RTI WDP TPC



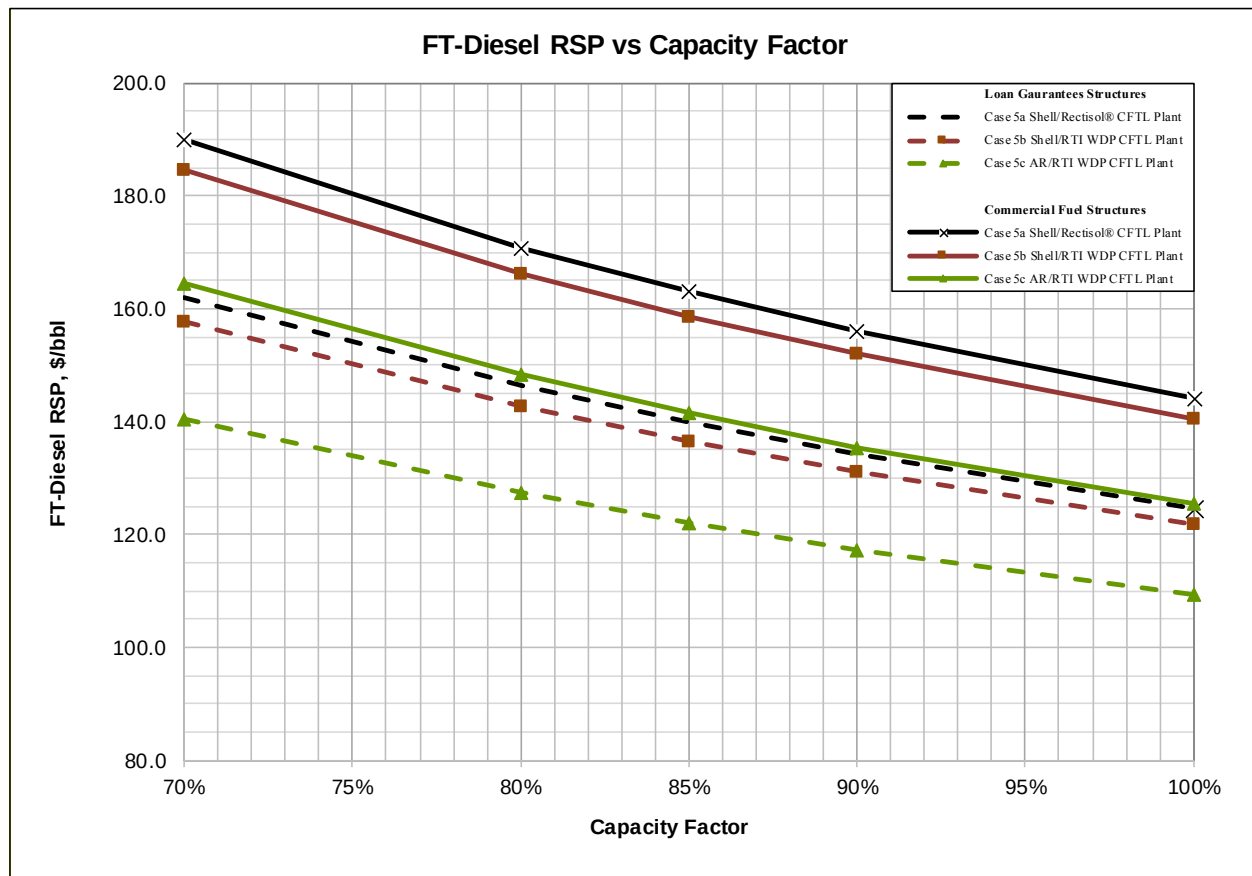
From Figure 7-2, it can be seen that at the very high end of the RTI WDP TPC (+30%), the FT-Diesel RSPs for Cases 5B and 5C are roughly comparable to the RSPs for the Case 5A reference plant but are lower than the reference case RSPs for all other WDP TPC sensitivity estimates.

Under the commercial fuels financing structure, every 7.5% increase in RTI WDP TPC for Case 5C (7.1% for Case 5B), equivalent to about \$49 MM, increases the FT-Diesel RSP by \$1/bbl. Under the loan guarantee financing structure, every 9.2% increase in RTI WDP TPC, equivalent to about \$61 MM, increases the FT-Diesel RSP by \$1/bbl.

7.3 CAPACITY FACTOR

The baseline CTL plant capacity factor used in this study is 90%. Figure 7-3 shows how the FT-Diesel product RSP varies with plant capacity factor as it varies from 70% to 100%. As can be seen, the RSPs for Cases 5B and 5C are always lower than for the reference Case 5A.

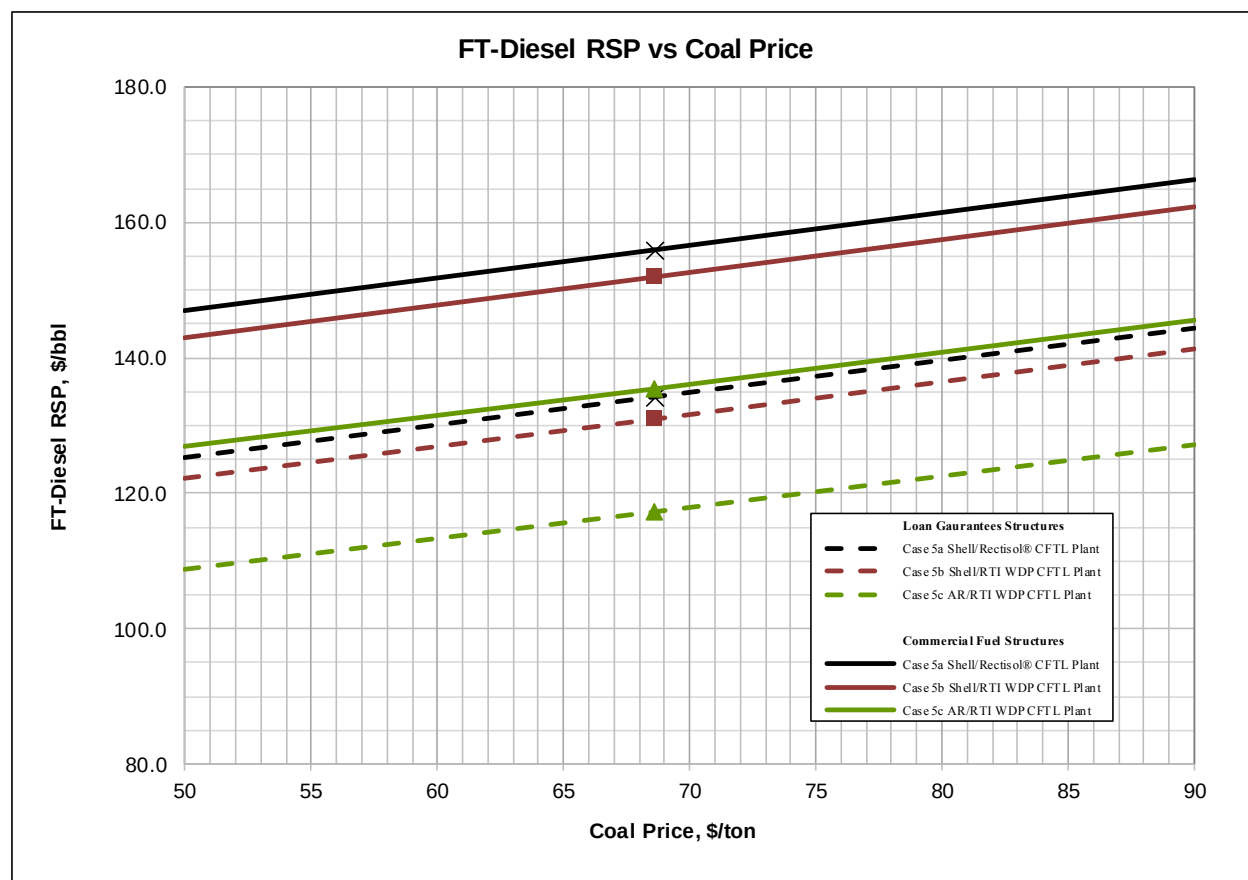
Figure 7-3
Sensitivity Analysis – RSP vs CTL Plant Capacity Factor



7.4 COAL PRICE

The Illinois #6 coal feedstock price used in this study was \$68.6/bbl, which was specified in the *QGESS Fuel Prices for Selected Feedstocks in NETL Studies* document. Figure 7-4 shows how the FT-Diesel RSP varies as the coal price is varied from \$50/bbl to \$90/ton.

Figure 7-4
Sensitivity Analysis – RSP vs Coal Feedstock Price



For the Shell gasifier cases (5A and 5B), the FT-Diesel RSPs increase at approximately the same rate with increases in coal prices since the feedstock requirements per unit of product are about the same. For these cases, the FT-Diesel RSP increases by \$1/bbl for every \$0.45/ton and \$0.52/ton increase in coal price, for the commercial and the loan guarantee financing structures respectively.

For the AR gasifier case 5C, the FT-Diesel RSPs also increase at about the same rate with increases in coal prices. However, the RSPs are less sensitive to coal pricing than the Shell gasifier cases since the AR gasifier has a higher cold gas efficiency. The FT-Diesel RSP for the AR gasifier cases increase by \$1/ton for every \$0.51/ton and \$0.59/ton of increase in coal price for the commercial fuels and loan guarantee financing structures respectively.

7.5 ELECTRICITY PRICE

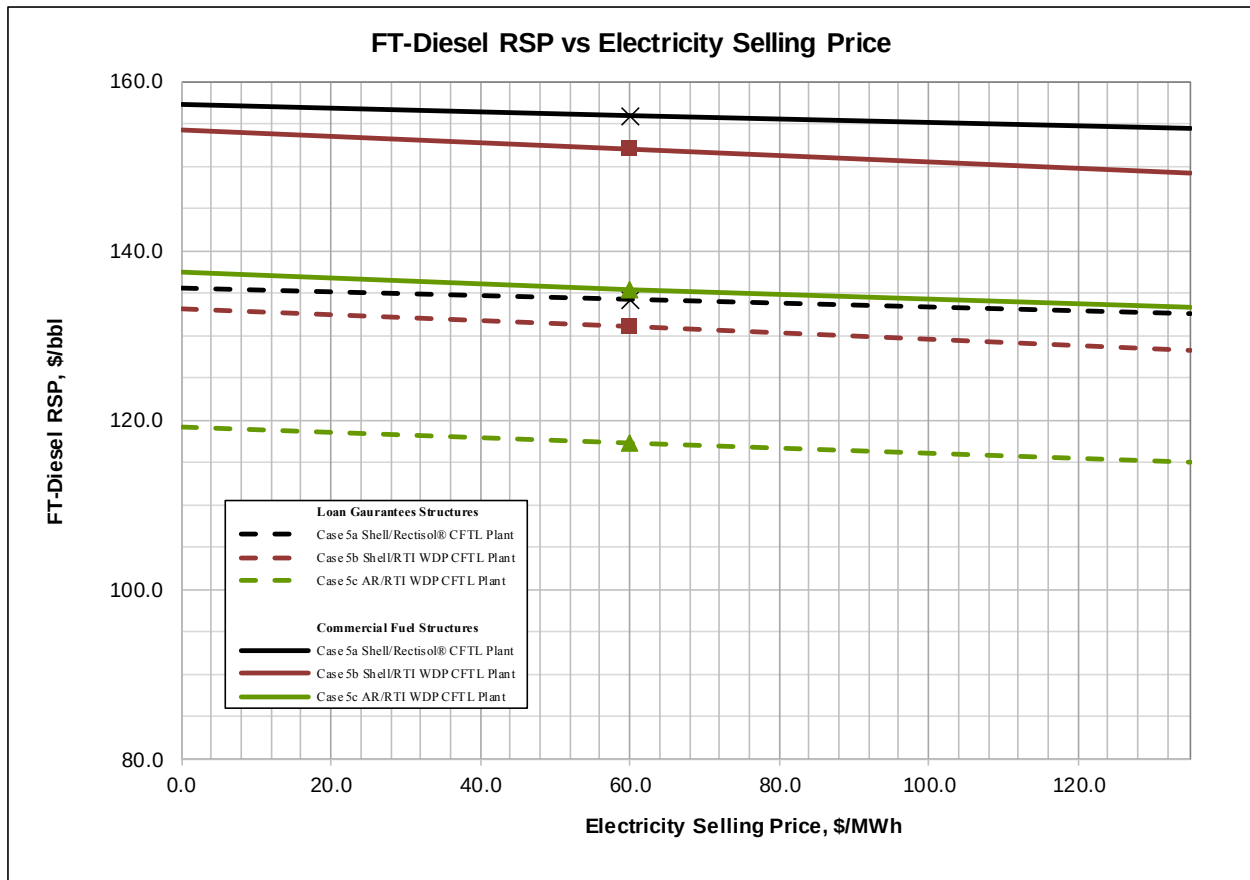
The combustion of FT fuel gas in the combustion turbine, combined with heat recovery in the HRSG and waste heat recovery in the CFTL plant from the exothermic FT-Diesel synthesis reaction and other processes, result in a substantial amount of excess electricity being generated in all three cases. The revenue derived from the sale of this excess electricity is applied to the cash flow and results in a lower RSP. The sensitivity of the RSP to the selling price of electricity is shown in Figure 7-6.

The electricity selling price used in this study was \$60.0/MWh. While the excess power can be sold to the grid, the sale may be at a steep discount as entities that are negotiating a power purchase agreement will know the power production is an inherent by-product of core FT-Diesel production operations. Consequently, the actual achieved transfer price for excess power would be highly project dependent. A range of \$0 - \$135/MWh is used in this sensitivity analysis.

The revenue derived from the sale of excess electricity accounts for 1 – 2 % of the total RSP for FT-diesel for all three cases, and is not a big driver for the FT-Diesel RSP.

Cases 5A using the Shell gasifier and conventional AGR process produce ~41 MW excess electricity whereas Case 5B and Case 5C which uses the Shell and AR gasifiers respectively, along with the RTI WDP for sulfur removal and AACRP for CO₂ removal, produces 58-70 MW of excess electricity. This is due to the use of the scrubber in Case 5A resulting in lower efficiency, and the use of the AFWGS in Cases 5B and 5C, leading to lower steam consumption in the water-gas shift process. Even though the AR gasifier has a higher cold efficiency, the excess fuel is used to generate additional FT liquids and does not impact the electricity production. Over the range of electricity pricing (0 - \$135/KWh) used for the sensitivity analysis, the RSP varies by ~1% for all three cases.

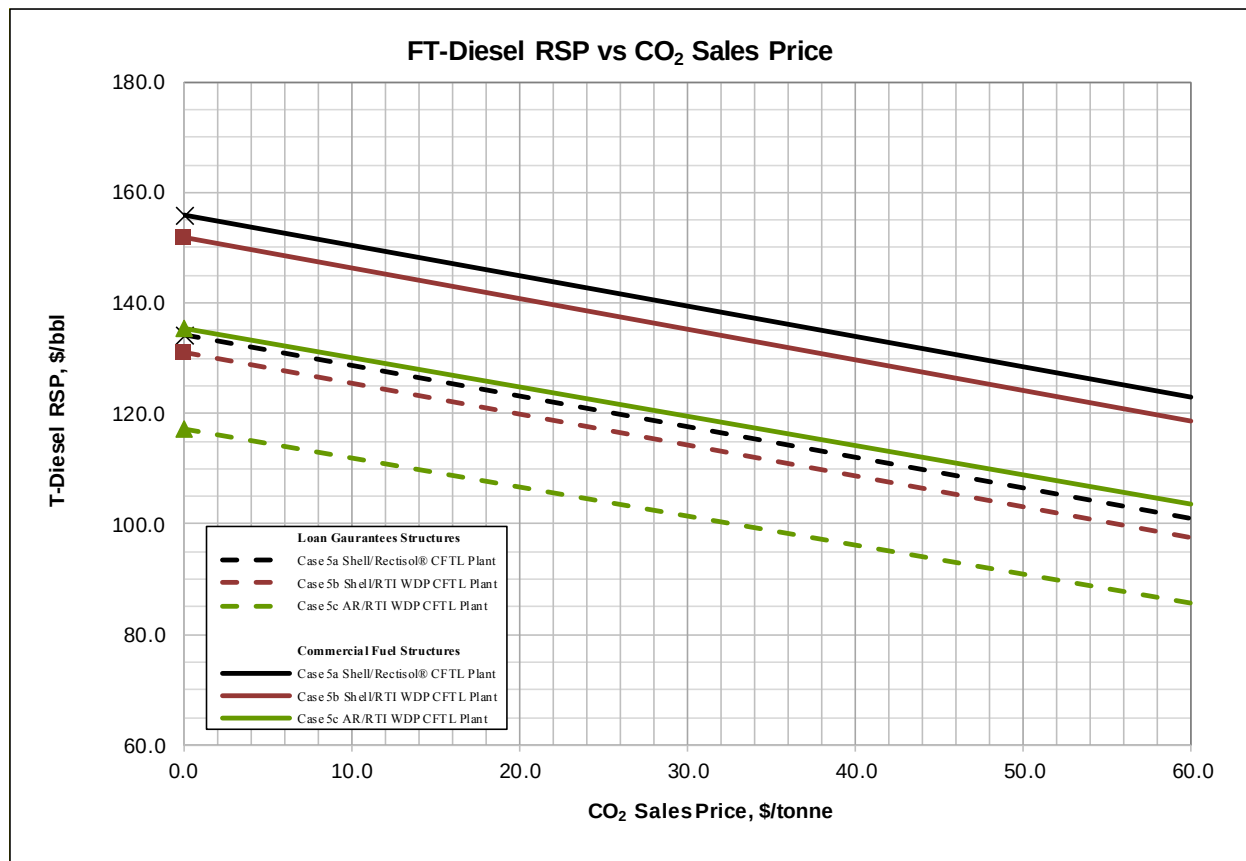
Figure 7-5
Sensitivity Analysis – RSP vs Electricity Price



7.6 CO₂ SALES PRICE

Options for carbon sequestration include both storage in a saline reservoir and usage in enhanced oil recovery (EOR). The impact of selling the captured CO₂ for EOR or other uses at various plant gate sales price, ranging from \$0-\$60/tonne, is shown in Figure 7-6. Sensitivity to CO₂ sales at plant gate prices is shown in Figure 7-6.

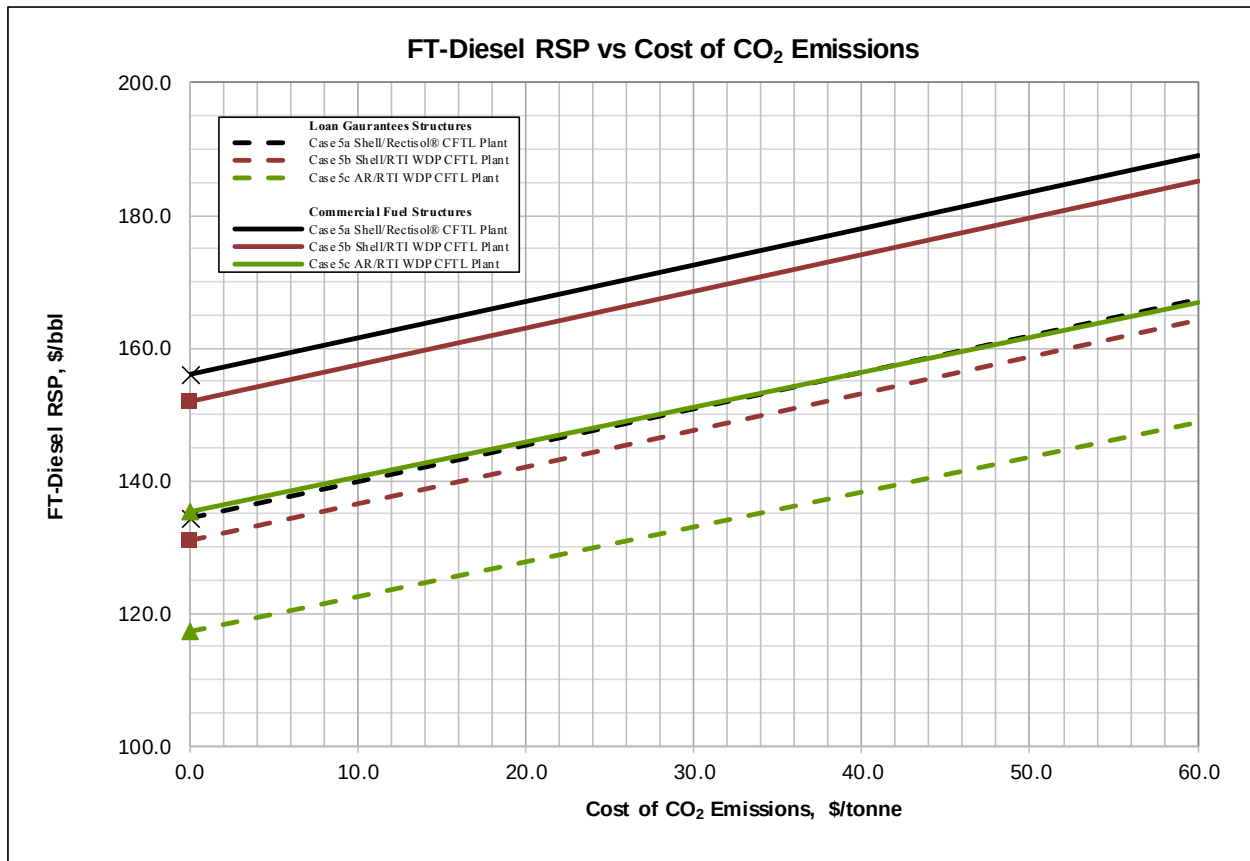
Figure 7-6
Sensitivity Analysis – RSP vs CO₂ Sales Price



7.7 COST OF CO₂ EMISSIONS

The sensitivity to CO₂ emissions costs is shown in Figure 7-8. The baseline case assumes that there are no costs associated with venting CO₂ to the atmosphere (\$0/tonne). The cost of CO₂ emissions is subsequently varied to a maximum of \$60/tonne to determine its effect on FT-Diesel RSP.

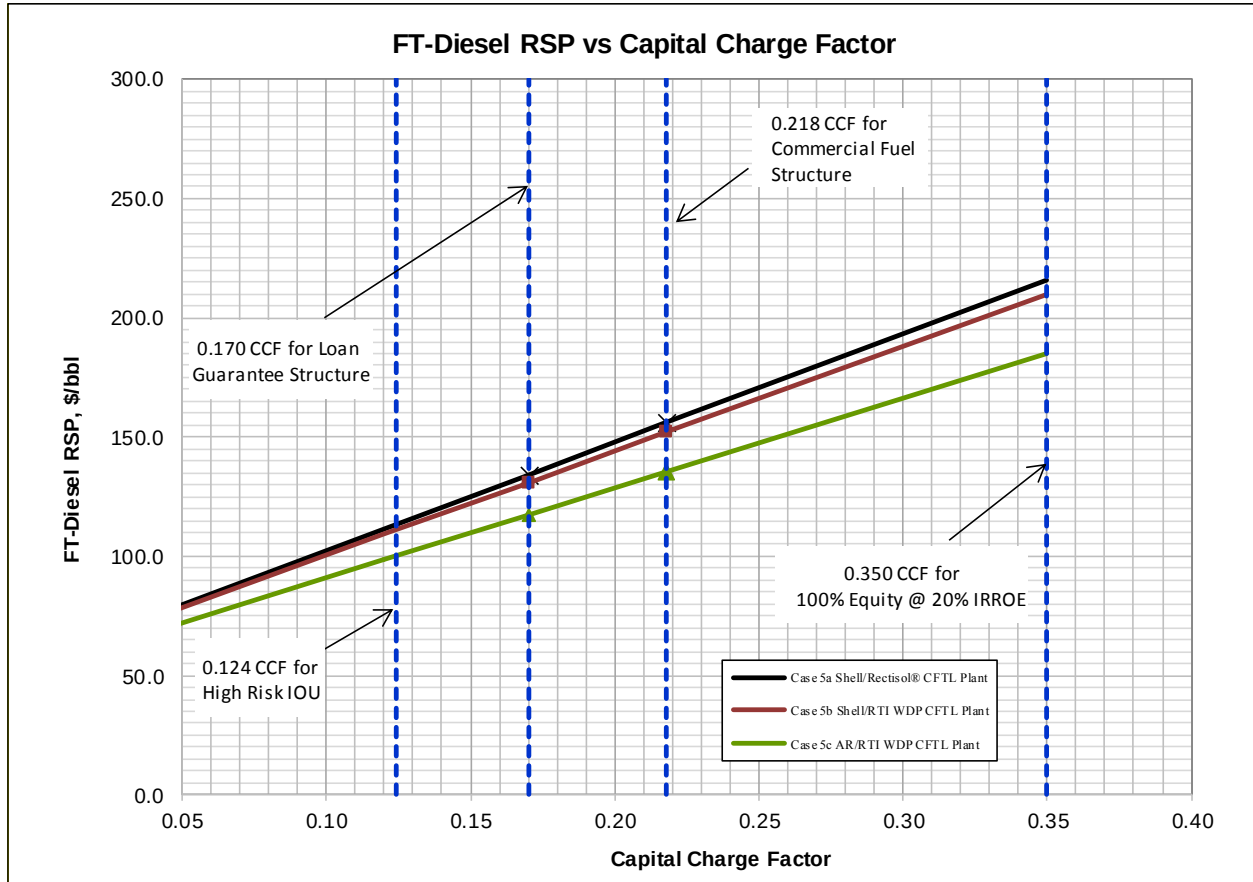
Figure 7-7
Sensitivity Analysis – RSP vs Cost of CO₂ Emissions



7.8 CAPITAL CHARGE FACTOR

Two financial structures were assumed for calculating FT-Diesel RSPs in this study as described in Section 2.14.1. These structures are based on typical values for fuel projects with and without loan guarantees or government subsidies. The assumed values were used in the PSFM to establish the CCF, which is the portion of the TOC to include in the annual cost of producing a product, for each financial structure. The sensitivity of the FT-Diesel RSP to the CCF is illustrated in Figure 7-9. The FT-Diesel RSP values were calculated for CCFs ranging from 0.10 to 0.35. The lines denoting CCFs of 0.124 (the value estimated for a high-risk investor-owned utility (IOU) project assuming 45% debt at 5.5% interest and 55% equity and 12% internal rate of return on equity (IRROE)) and 0.35 (the value estimated for 100% equity and 20% IRROE) are also included in the chart.

Figure 7-8
Sensitivity Analysis – RSP vs CCF



8.1 STUDY OBJECTIVE

The objective of this study was to assess the benefits of integrating the technologies offered by RTI and AR for reducing the capital and production costs for syngas generation and cleanup in a CTL plant with carbon capture

The reference case for this study uses a Shell gasifier to convert Illinois #6 coal into FT liquids using a water gas shift reactor to alter the H₂/CO ratio of the raw syngas, a Rectisol[®] syngas cleanup process, a cobalt-based FT process with product upgrading to produce the final fuel products, and a combined cycle system to recover the auxiliary power from the hydrocarbon byproduct. This case was developed by combining the information from different DOE studies on CTL using a Shell gasifier with an iron-based catalyst, and a GTL study using a cobalt-based catalyst.

To fully evaluate the benefits of integrating these technologies, a total of three case studies were completed. The specific technologies included in each of the three CFTL plant configurations are identified in the CFTL case study matrix shown in Table 8-1.

Table 8-1
Case Study Matrix for CFTL-Liquids Production with CO₂ Capture

	Case 5A ¹	Case 5B	Case 5C
Gasification Technology			
Shell Gasification	✓	✓	
AR Gasification Technology			✓
Gas Cleanup			
Rectisol [®] for CO ₂ and Sulfur Removal ²	✓		
RTI WDP with Activated Amine CO ₂ Recovery Process (AACRP)		✓	✓
Water Gas Shift			
Sour Shift	✓		
Advanced Fixed-Bed Water-Gas Shift		✓	✓
FT Synthesis			
Conventional technology ³	✓	✓	✓
CO ₂ Drying and Compression (to 2,215 psia)	✓	✓	✓

¹ Case 5A is the reference case. Gasifier selection depends on selected DOE NETL baseline studies.

² Rectisol[®] remove sulfur and CO₂. Selection and any additional trace contaminant cleanup technologies needed will be as defined by DOE NETL baseline studies.

³ Baseline FT technology will be established based on DOE NETL guidelines/recommendations

8.2 RESULTS SUMMARY

Table 8-2 shows a summary comparison of the CAPEX, OPEX, FT liquids production, plant power import/export, auxiliary power consumption, FT-products RSP, and cost of CO₂ captured/avoided for the three cases.

Table 8-2
CTL Results Summary

Case	Case 5A		Case 5B		Case 5C	
Gasifier	Shell		Shell		Aerojet	
Sulfur and CO ₂ Removal	Dual Stage Rectisol [®]		RTI WDP/AACRP		RTI WDP/AACRP	
CAPEX, \$MM						
Total Installed Cost (TIC)	\$3,882		\$3,734		\$3,220	
Total Plant Cost (TPC)	\$5,405		\$5,187		\$4,577	
Total Overnight Cost (TOC)	\$6,771		\$6,530		\$5,797	
OPEX, \$MM/yr (100% Capacity Factor Basis)						
Fixed Operating Cost (OC _{fix})	\$205.8		\$198.9		\$179.8	
Variable Operating Cost, less Fuel (OC _{var})	\$176.1		\$190.0		\$179.3	
Coal Fuel (OC _{fuel})	\$526.0		\$526.0		\$526.0	
Import/(Export) Power (OC _{Power})	(\$21.7)		(\$36.9)		(\$30.7)	
Total OPEX	\$886.1		\$878.0		\$854.3	
Plant Output						
Crude FT Diesel Product, bbl/day	34,992		34,994		36,115	
Crude FT Naphtha Product, bbl/day	15,008		15,006		15,486	
Net Power Output, MWe	41.3		70.2		58.4	
Power Generation						
Gas Turbine Power	122.5		101.5		104.9	
Steam Turbine Power	403.2		448.6		414.2	
Syngas Turboexpansion Power	7.3		10.0		11.4	
Total Gross Power	533.0		560.0		530.4	
Net Power Generation/(Consumption), MWe	41.3		70.2		58.4	
Common Factors Used for RSP Calculation						
Capital Charge Factor (CCF)	0.1700	0.218	0.1700	0.218	0.1700	0.218
Capacity Factor (CF)	0.9		0.9		0.9	
Required Selling Price^{A,B,C}						
RSP FT Diesel, \$/bbl _{FTD}	134.3	156.0	131.0	152.0	117.3	135.6
RSP FT Naphtha, \$/bbl _{FTN}	94.2	109.2	89.9	104.6	80.4	93.1
Equivalent Crude Oil Price, \$/bbl _{ECO}	116.5	135.4	113.7	131.9	101.8	117.7
Equivalent Petroleum Diesel Price, \$/bbl _{EPD}	145.7	169.3	142.1	164.9	127.3	147.1
F-T Diesel Required Selling Price Breakdown						
CAPEX	77.02	98.76	74.27	95.25	63.89	81.93
Fixed O&M	13.77	13.77	13.31	13.31	11.65	11.65
Variable O&M	10.60	10.60	11.44	11.44	10.46	10.46
Fuel	31.67	31.67	31.67	31.67	30.69	30.69
Power Generation	-1.31	-1.31	-2.22	-2.22	-1.79	-1.79
CO ₂ T&S	2.50	2.50	2.52	2.52	2.39	2.39
Total	134.25	156.00	131.00	151.97	117.29	135.33
CO₂ Capture						
CO ₂ Capture Product, tpd	25,116		25,280		24,729	
Carbon Captured (as % Carbon in Fuel)	92.5%		92.8%		92.6%	
Costs of CO₂ Captured^{A,B,C} (\$/ton)	13.92	15.00	13.44	14.49	13.47	15.04

^A Capacity factor assumed to be 90 percent

^B Values shown are for two financial structures

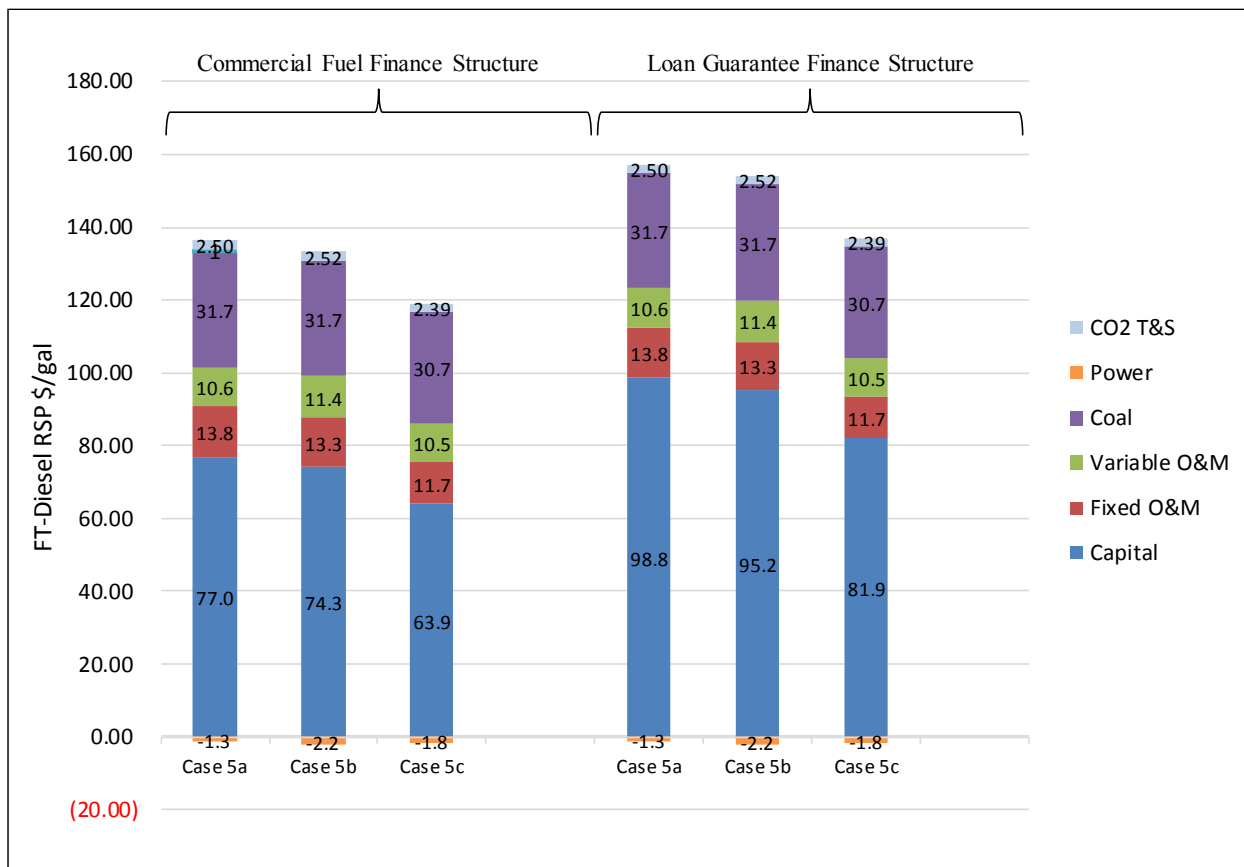
The first (lower value) is based on the loan guarantee finance structure

The second (higher value) is based on the commercial fuels finance structure

^C Includes CO₂ T&S

Figure 8-1 illustrates the breakdown of the cost of production of FT-diesel. Since the results for the RSP of FT-naphtha will be similar due to the fixed ratio of naphtha and diesel produced, the results are shown for only the FT-diesel and similar conclusions apply to the FT-naphtha production. It can be seen from Figure 8-1 that the cases differ most significantly in terms of capital cost on a \$/bbl basis.

Figure 8-1 Components of FT Diesel



This breakdown by component of the FT-diesel RSP for the different cases illustrates the relative contribution of each component. In all cases, the capital cost accounts for 54-63% of the FT-diesel RSP. Both the AR gasification and RTI advanced warm syngas cleanup significantly reduce capital cost and FT diesel RSP. The AR gasification system reduces the capital cost of the overall gasification block (including the dry feed system) by 35% and results in reduction in TPC of the entire CFTL plant by about \$610 MM or a 11.5% decrease. The RTI advanced warm syngas cleanup system reduces capital costs of the overall gas cleanup block (including AGR, SRU, LTGC, and WGS) by 24% and results in reduction in TPC of the entire CFTL plant by \$211 MM or about 4%. These capital cost reductions result in a 10% and 3% decrease in the FT diesel RSP for the AR gasification and RTI advanced warm syngas cleanup technologies respectively, for both financial structures. The combined advanced technologies yield a 15% reduction in overall CFTL capital cost and a 13% lower RSP of FT diesel liquids compared to the reference case.

In Case 5A, the CTL plant has a total of 9 Shell gasifiers with 8 in continuous operation and one spare. Based on the coal throughput, the size of these Shell gasifiers is about 2,000 tpd. Based on the assumption that the reference case should reflect actual commercially operating plants, this gasifier selection is practical. However, this study focuses on evaluating emerging technologies which are expected to be demonstrated commercially in the next 5 years. Because of this approach, the AR gasification system included a 3,600 tpd coal gasifier. Recently Shell announced that they are constructing a CTL plant in China with Shell gasifiers with a 3200 tpd coal intake capacity. [19] If Shell's larger gasifier had been used in the reference case, a total of 6 continuously operating gasifiers would have been required, which is identical to the number of AR gasifiers used in this study. There is no doubt that use of the AR gasification system would still result in a substantial reduction in capital cost (because the case studies in this project have consistently shown that AR gasifiers cost substantially less than equivalent-sized Shell gasifiers), but the total magnitude of any cost reduction would be somewhat smaller.

The gas turbine (GT) power generated by combusting the FT fuel gas is quite different for the three cases. Case 5A produces 122.5 MW of power in the GT turbine. Compared to Case 5A, Case 5B and 5C produce 101.5 MW and 105 MW of GT power respectively. This difference is due to the reduced hydrogen content in the FT fuel gas due to the hydrogen that is diverted for reducing the SO₂ stream in the DSRP for Cases 5B and 5C. Case 5C uses the AR gasifier which has a higher cold gas efficiency. The benefit from the increased efficiency is utilized by producing excess FT liquids rather than excess power, since the FT liquid is a higher value product.

The steam turbine (ST) power produced in the three cases are also quite different. The base case, Case 5A, produces 403 MW of power, compared to 449 MW in Case 5B and 414 MW in Case 5C. Compared to Case 5A, ST power in Case 5B is higher due to the net excess steam production in Case 5B. Case 5B uses the RTI WDP for warm-gas desulfurization. As such, the syngas in Case 5B, instead of being sent to the scrubber, is sent to the RTI WDP process at 750°F, and the gas exiting the RTI WDP is used to raise additional high pressure steam. Also, the AFWGS process uses less steam compared to the sour water gas shift in Case 5A. Both choices result in a larger amount of HP steam that is available for power production in the ST.

Compared to case 5B, Case 5C results in a lower ST power production, even though both processes use the RTI WDP process and the advanced HT shift. This difference in ST power results from the quenching of the syngas that occurs in the AR gasifier compared to the cooling of the syngas in the Shell gasifier by raising HP steam in the radiant and duct coolers of the Shell gasifier.

There is also 7-11 MW of power that is generated by the expansion of the syngas before sending it to the FT synthesis reactor.

Case 5A results in 41 MW of net power that is exported. Improved steam generation from process heat recovery and reduced steam consumption in the AFWGS, Case 5B and Case 5C results in 70 MW and 58 MW of net power, respectively.

The ST power generation in Case 5A is ~400 MW. This is higher than the ST power generation in the base case from the DOE CTL study [2], which is 315 MW. The two cases differ in the extent of the shift reaction prior to the FT synthesis and the choice of the FT synthesis catalyst. The DOE base case uses the unshifted syngas out of the gasifier at a H_2/CO ratio of 0.73 and uses a Fe-catalyst whereas Case 5A shifts the syngas to achieve H_2/CO ratio of 2.0-2.1, and uses the Co-catalyst. The differences in the power generation is probably due to better heat recovery through HP steam production in the water gas shift area in Case 5A compared to generation of medium pressure 360 psig steam in the FT area in the DOE study base case.

The variable OPEX costs are dominated by the costs for the FT catalyst, accounting for 18-19% of the total variable operating costs. The initial fill for the FT catalyst was scaled from the Bechtel study using a factored scaling approach. The total initial fill determined was found to be roughly 4 times higher than the catalyst fill reported in the DOE CTL [2] and GTL [3] studies. Since the same assumptions about cost and catalyst annual consumption were used on these three cases, any benefits or differences noted for these three cases are based on actual differences rather than differences in assumptions. The annual consumption rate for the catalyst was determined from the reported lifetime of 2.74 years for the Co-catalyst.

The cost of the FT catalyst was assumed to be \$20/lb. This is similar to other Co-based catalysts. Both the DOE CTL and the GTL report uses a cost of \$7.15/lb for the FT catalyst, even though the CTL study uses a Fe-catalyst and the GTL study uses a Co-catalyst.

The reference case (Case 5A) and thus comparison Cases 5B and 5C were based on use of Illinois #6 coal, which has a high chlorine content (0.29 wt%) and produces a raw syngas with 900 ppmv of HCl. Since high levels of HCl in the syngas are detrimental to all downstream chemical processes, this required a process for removal of the HCl. In Case 5A, the chlorine is removed in the syngas scrubber and the chlorine is treated as part of the sour water processing. Case 5B and Case 5C use a warm-temperature non-regenerable HCl guard bed process to remove the HCl from the syngas upstream of the RTI WDP process, which improves the overall thermal efficiency of the process as described previously.

Although the technical aspects of the HCl removal could be dealt with as part of these case studies, effective costing of the HCl removal was more challenging. Sufficient details were not provided in the DOE reference cases to isolate the chlorine treatment cost from the sour gas treatment costs. As all three cases had sour water processing, the factored estimating approach used included the chlorine treatment cost, which was required in Case 5A, but was not needed in Cases 5B and 5C where the chlorine was captured by a warm-temperature non-regenerable HCl sorbent. Replacement of this HCl guard bed sorbent is about \$7MM in the annual operating costs.

If the variable cost of chlorine treatment in the sour water processing is assumed to be equivalent to the variable cost for replacement of the HCl sorbent, the cost of chlorine removal is about 5% of the total variable annual operating cost. Since the variable operating expenses only contribute to ~10.5% of the RSP, the cost of the HCl guard bed contributes only ~0.5% of the total RSP. Therefore, the overestimated cost of chlorine removal for Cases 5B and 5C is about 5% of the total annual variable costs and thus would be expected to result in about a 0.5% or \$0.70

overestimation of diesel RSP. However, a more detailed analysis of these costs should be conducted to effectively understand the cost of chlorine removal for high chlorine coals for these advanced technologies.

For most other coals, which do not contain such high levels of chlorine, the use of HCl guard beds would not be required and their resultant capital and operating costs would thus be avoided, which would increase the relative economic benefits for Cases 5B and 5C reported above.

A key criterion for the CFTL plants in this study was to achieve near-zero levels of emissions. Because of the sensitivity of the cobalt-based FT catalyst used in the FT synthesis process, the gas phase sulfur emissions were effectively reduced to zero. Although other contaminants were not tracked as meticulously as sulfur, their gas phase emissions were also essentially reduced to zero because of the syngas cleaning required for the cobalt-based FT catalyst.

Because the byproduct hydrocarbons from the FT synthesis and upgrading processes are used for fuel gas and to generate electricity, there are some net CO₂ gas phase emissions. However, all three cases result in between 92% and 93% overall carbon capture, which is because CO₂ capture is an inherent part of the FT process requirements. Consequently, the cost of CO₂ captured ranges from \$13.4- \$13.9 per barrel of CO₂ for all three cases for the loan guarantee finance structure, because the only required additional storage and sequestration cost comes from compression of the captured CO₂ to 2215 psia for sequestration. Interestingly, the CO₂ footprint for the diesel and naphtha products is about 147 lb of CO₂/bbl for Case 5A and 143 lb CO₂/bbl for Cases 5B and 5C. With all the net CO₂ emissions being assigned to the FT products, the electricity being generated has a zero-carbon footprint. Because the electricity can be viewed as an inherent byproduct of the CFTL plant, market forces will attempt to drive down the price of this electricity. Because this electricity has essentially no carbon footprint, it is possible that it could be negotiated to command a premium price in the proper regulatory environment.

The CFTL plants achieve lower overall gas phase emissions than either the IGCC or CTM plants also studied under this project. But even for these CFTL plants, the advanced technologies, and in particular RTI's warm syngas cleanup technologies, result in overall emissions reductions versus the reference case.

The substantial benefits from integration of AR's gasification and RTI's advanced syngas cleanup technologies into a CFTL plant, which include significant capital cost reductions, lower emissions, and improved energy efficiency, will generate market forces that will drive the implementation of the AR and RTI technologies to realize these benefits. The sensitivity analysis completed on the impact of increasing the estimated capital costs of the AR gasification and the RTI advanced syngas cleanup technologies by as much as 30% still demonstrated significantly lower FT-diesel RSPs. All these factors indicate that when ready for commercial deployment, these technologies will be favored over existing commercial technologies.

Appendix A

Acronyms and Abbreviations

°F	Degree Fahrenheit
AACRP	Activated Amine CO ₂ Recovery Process
AFWGS	Advanced Fixed-Bed Water Gas Shift
AGR	Acid Gas Removal
AOI	Area of Interest
AR	Aerojet Rocketdyne
Ar	Argon
ASU	Air Separation Unit
B/L	Battery Limit
BACT	Best Available Control Technology
BEC	Bare Erected Cost
BFD	Block Flow Diagram
BFW	Boiler Feed Water
BOP	Balance of Plant
Btu	British Thermal Unit
CAPEX	Capital Expenditure
CCF	Capital Charge Factor
CF	Capacity Factor
CH ₄	Methane
Circ	Circulating
CMT	Constant Maturity Treasury
Co	Cobalt
CO	Carbon Monoxide
CO ₂	Carbon Dioxide
COE	Cost of Electricity
COS	Carbonyl Sulfide
CFTL	Coal-to-Fischer-Tropsch-Liquids
CTL	Coal-to-liquids
CTM	Coal-to-Methanol
CTG	Combustion Turbine Power Generation
CW	Cooling Water
CWP	Circulating Water Pump
DBT	Dry Bulb Temperature
DCS	Distributed Control System
DOE	U.S. Department of Energy
DSP	Dry Solids Pump
DSRP	Direct Sulfur Reduction Process

EPC	Engineering, Procurement and Construction
EPRI	Electric Power Research Institute
FCC	Fluid Catalytic Cracking
Fe	Iron
FO	Fuel Oil
FOA	Funding Opportunity Announcement
FT	Fischer-Tropsch
ft	feet
gal	Gallon
GE	General Electric
GT	Gas Turbine
H ₂	Hydrogen
H ₂ O	Water
H ₂ S	Hydrogen Sulfide
Hg	Mercury
HGCU	Hot Gas Clean Up
HHV	Higher Heating Value
HMB	Heat and Material Balance
HP	High Pressure
hr	Hour
HRSR	Heat Recovery Steam Generator
I & C	Instrumentation & Control
IGCC	Integrated Gasification Combined Cycle
IOU	Investor Owned Utility
IP	Intermediate Pressure
kWe	Kilowatt electric
kWh	kilowatt hour
LAER	Lowest available emissions rate
lb	Pound Mass
LHV	Lower Heating Value
LIBOR	London Interbank Offered Rates
LNB	Low NO _x Burner
LP	Low Pressure
LTFT	Low Temperature Fischer-Tropsch
LTGC	Low temperature gas cooling
max	Maximum
MDEA	Methyldiethanolamine
ME	Major Equipment
MEC	Major Equipment Cost
MeOH	Methanol

min	Minimum
Misc	Miscellaneous
MM	million
MU	Makeup
MWe	Megawatt electric
MWh	megawatt hour
N ₂	Nitrogen
NETL	National Energy Technology Laboratory
NGCC	Natural Gas Combined Cycle
NO _x	Nitrogen oxides
NSR	New Source Review
O&M	Operating and Maintenance
O ₂	Oxygen
OPEX	Operating Expenditure
OSBL	Outside Battery Limit
PC	Pulverized Coal
PCC	Post-Combustion Capture
PFB	Pressurized Feed Bin
PFD	Process Flow Diagram
PM	Particulate Matter
ppmv	Parts per Million by Volume
ppmW, ppmw	Parts per Million by Weight
PRB	Powder River Basin
PSFM	Power Systems Financial Model
psi	Pounds Per Square Inch
psia	Pounds Per Square Inch, absolute
psig	Pounds Per Square Inch, gauge
QGESS	Quality Guidelines for Energy System Studies
RTI	Research Triangle Institute or RTI International
RSP	Required Selling Price
SC	Supercritical
SCGP	Shell Coal Gasification Process
scfm	Standard Cubic Feet per Minute
SO ₂	Sulfur Dioxide
SOPO	Statement of Project Objectives
SRU	Sulfur Recovery Unit
ST	Steam Turbine
STG	Steam Turbine Power Generation
SWGS	Sour water gas shift
T&S	Transportation and Storage

TDC	Total Direct Cost
TEA	Techno-Economic Analysis
TFC	Total Field Cost
TG	Turbine Generator
TGTU	Tail Gas Treatment Unit
TIC	Total Installed Cost
TOC	Total Overnight Cost
TPC	Total Plant Cost
tpd	tons per day
US, USA	United States of America
vol%	Percentage by Volume
WBT	Wet Bulb Temperature
WDP	Warm Gas Desulfurization Process
WGS	Water Gas Shift
WHR	Waste Heat Recovery
WT	Waste Treatment
ZnO	Zinc oxide

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