FINAL SCIENTIFIC/TECHNICAL REPORT

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Hybrid Molten Bed Gasifier for High Hydrogen Syngas Production

SUBMITTED BY

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DISCLAIMER

ABSTRACT

The Gas Technology Institute (GTI) has developed a hybrid Molten Bed (HMB) dual coal-natural gas fueled gasification process. In the HMB gasifier gas is fired under partial oxidation conditions with oxygen into a bed of molten coal slag which produces the heat and steam needed to drive the endothermic gasification of coal charged to the molten bed. The gasification process is made more efficient than other gasifiers by recuperating heat from its walls and from the hot, raw syngas through endothermic steam-methane reforming of the natural gas. Chemical energy is returned as fuel to the gasifier. Control of independent variables, including coal and natural gas feed rates, and the oxygen to natural gas ratio enables the HMB gasifier to produce syngas with a controlled H_2/CO ratio of 1 to >6. The syngas composition can be optimized for producing electricity by integrated gasification combined cycle (IGCC) or liquid fuels or chemicals by Fischer-Tropsch (FT) or other catalytic processes.

Project partner Nexant, Inc. carried out techno-economic analyses (TEA) of the HMB process. Comparisons were made with the Shell entrained flow gasifier as a base case. The first TEA considered HMB gasification for IGCC power production. HMB gasification was found to have better overall IGCC economics as compared to the coal only Shell gasification process. For non-Recuperative Coal Feed only Operation (Case 1), no efficiency difference was found vs Shell IGCC - 31.2% vs 31.2%, but the cost of the HMB gasifier is lower vs Shell ($400 MM vs $750 MM), and the cost of electricity (COE) is lower COE (122.8 mills/kWh vs 144.8 mills/kWh). For recuperative operation with coal-natural gas co-feed and no reformer (Case 2), the efficiency improvement was found to be 2.2% (33.4% vs. 31.2%) from steam and natural gas preheats, but there was found to be a higher COE (125.0 mills/kWh vs 122.8 mills/kWh) due to higher fuel cost. For recuperative operation with coal-natural gas co-feed with an external reformer (Case 3) an additional efficiency improvement of 0.8% was found vs Case 2 (35.2% vs. 33.4%) by heat recuperation through an external steam reformer and COE was found to be higher (125.8 mills/kWh vs 125.0 mills/kWh for Case 2) due to the added cost of the steam reformer.

Nexant carried out techno-economic analyses (TEA) of HMB gasification is Fischer-Tropsch mode compared with the Shell entrained flow gasifier as a base case. Cases considered assumed a 55/45 coal/natural gas blend. The first case assumed direct raw gas reforming. The second gas considered parallel indirect reforming, and the third case assumed series indirect reforming. In all HMB FT cases, the cost of power (COP) based on diesel production was equal to or 1 to 2% higher than the Shell baseline case. The HMB cases all had lower fixed and variable operating costs compared with the Shell case, but they all had significantly higher fuel costs because gas has a higher price than powder river basin coal.

Laboratory HMB testing was carried out in a single-burner unit built in the GTI Industrial Combustion Laboratory. A series of tests were carried out with 200 mesh Illinois #6 coal and either natural gas or one of two syngas mixtures. Steam was added in the syngas and a portion of the natural gas tests. Coal feed rate was 20 to 36 lb/h and represented 49 to 57% of total energy input. Oxygen to gas stoichiometric ratio was between 1.6 and 2.1. Gases and oxygen were introduced into the 375 lb molten bed through the vertically mounted floor burner. Coal was delivered from a calibrated feeder by pneumatic nitrogen transport. Testing with natural gas and no steam found that H2/CO ratios of product syngas were 0.52 to 0.60. Product syngas quality

measured as higher H2/CO ratio improved as the O2 to fuel ratio was decreased and as the fraction of fuel as coal increased. Increasing coal from 49% to 56% increased H2/CO and the percent CO in product syngas. Adding feed steam and switching feed natural gas to syngas both increased product H2/CO (to as high as 0.78) with small impact on the percent CO in product syngas. A longer coal feed tube allowing coal introduction just above the molten bed also increased H2/CO ratio. HMB test operations were highly stable. Bed and product syngas temperatures were stable at all test points. The burner operated flawlessly over the full range of oxygen to fuel gas ratio, with different fuel gases, and with steam.

Experiments did not generate syngas with the desired H2/CO ratio of 1.5 to 2, but results were encouraging. Operation at higher temperature and with a lower oxygen to fuel gas ratio are expected to be needed to achieve the desired H2/CO ratio of 2.0.

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EXECUTIVE SUMMARY

The techno-economic analyses of the hybrid molten bed gasification technology and laboratory testing of the HMB process were carried out in this project by the Gas Technology Institute and partner Nexant, Inc. under contract with the US Department of Energy's National Energy Technology Laboratory. This report includes the results of two complete IGCC and Fischer-Tropsch TEA analyses comparing HMB gasification with the Shell slagging gasification process as a base case. Also included are the results of the laboratory simulation tests of the HMB process using Illinois #6 coal fed along with natural gas, two different syngases, and steam.

Work in this 18-month project was carried out in three main Tasks. Task 2 was completed first and involved modeling, mass and energy balances, and gasification process design. The results of this work were provided to Nexant as input to the TEA IGCC and FT configurations studied in detail in Task 3. The results of Task 2 were also used to guide the design of the laboratory-scale testing of the HMB concept in the submerged combustion melting test facility in GTI's industrial combustion laboratory. All project work was completed on time and budget. A project close-out meeting reviewing project results was conducted on April 1, 2015 at GTI in Des Plaines, IL.

The hybrid molten bed gasification process techno-economic analyses found that the HMB process is both technically and economically attractive compared with the Shell entrained flow gasification process. In IGCC configuration, HMB gasification provides both efficiency and cost benefits. In Fischer-Tropsch configuration, HMB shows small benefits, primarily because even at current low natural gas prices, natural gas is more expensive than coal on an energy cost basis. HMB gasification was found in the TEA to improve the overall IGCC economics as compared to the coal only Shell gasification process.

Operationally, the HMB process proved to be robust and easy to operate. The burner was stable over the full oxygen to fuel firing range (0.8 to 1.05 of fuel gas stoichiometry) and with all fuel gases (natural gas and two syngas compositions), with steam, and without steam. The lower Btu content of the syngases presented no combustion difficulties.

The molten bed was stable throughout testing. The molten bed was easily established as a bed of molten glass. As the composition changed from glass cullet to cullet with slag, no instabilities were encountered. The bed temperature and product syngas temperature remained stable throughout testing, demonstrating that the bed serves as a good heat sink for the gasification process. Product syngas temperature measured above the bed was stable at ~1600ºF.

Testing found that syngas quality measured as H2/CO ratio increased with decreasing oxygen to fuel gas stoichiometric ratio, higher steam to inlet carbon ratio, higher temperature, and syngas compared with natural gas. The highest H2/CO ratios achieved were in the range of 0.70 to 0.78. These values are well below the targets of 1.5 to 2.0 that were expected and were predicted by modeling. The team, however, is encouraged that the HMB process can and will achieve H2/CO ratios up to 2.0. Changes needed include direct injection of coal into the molten bed of slag to prevent coal particle bypass into the product gas stream, elevation of the molten bed temperature

to approximately 2500ºF, and further decrease of the oxygen to fuel gas ratio to well below the 0.85 minimum ratio used in the testing in this project.

REPORT DETAILS

Objectives

The project objective was to conduct HMB gasification process techno-economic analyses and laboratory process tests. The techno-economic analyses (plant efficiency, cost of products, environmental performance) were carried out to maximize IGCC power production in one configuration and to maximize Fischer-Tropsch (FT) diesel yield in a second configuration. Baseline and parametric analyses were conducted for both power and diesel production with carbon capture in cases including a conventional slagging gasifier and the HMB gasifier. Laboratory HMB process tests were conducted at GTI to collect data to support the techno-economic analyses carried out by Nexant and to guide process scale-up calculations.

Introduction and Background

Decades of effort by engineers and chemists have led to the development of a number of gasification technologies including GTI's U-GAS and HYGAS, Lurgi, Siemens, E-Gas, Shell, and many others. These technologies rely on fluidized beds, circulating fluidized beds, entrained flow, and slagging configurations. All approaches produce syngas with a H_2/CO ratio of 1.0 or lower because coal is rich in carbon and poor in hydrogen. Syngas is a valuable product as a fuel and can be even more valuable when used to produce electricity in an integrated gasification combined cycle (IGCC) plant, liquid products such as diesel and naphtha by Fischer-Tropsch (FT) catalysis, gasoline by Halder Topso or other liquid hydrocarbon synthesis (LHS) catalysis, methanol, or other chemicals. Coal syngas is not well suited for these uses and must be upgraded to a higher H_2/CO ratio (such as 2.1 for FT and as high as possible for IGCC) by often costly means. The most common means of upgrading syngas are by enhanced steam or catalytic steam gasification in which coal or other fuel is expended to generate steam needed in the gasifier and the water gas shift (WGS) reactor. IGCC or LHS gasification plants with carbon capture for sequestration are complex, so engineers strive to optimize the gasifier, but more importantly work to optimize efficiency, flexibility, and cost of the entire plant. The gasifier must be simple and flexible so overall plant cost can be minimized.

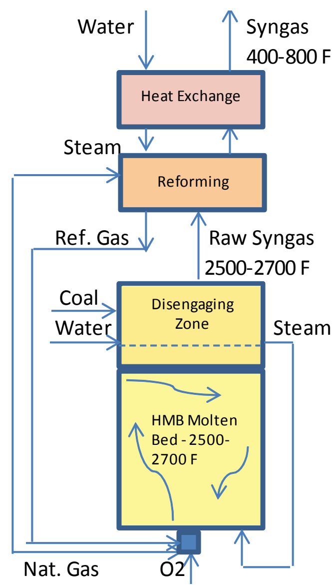


Figure 1. HMB

A second way to increase syngas H_2/CO ratio is to co-fire a gasifier with coal and a more hydrogen-rich fuel. Solid fuels such as biomass and lignin are good candidates, and co-feeding them with coal yields a more desirable syngas. But these fuels present challenges including solids handling complexities because they have lower energy density and differ in consistency from coal. They may not be available in large and regular quantities, and they may be unable alone to yield the syngas to attain the desired H_2/CO ratios making some steam addition still necessary. A more promising dual-feeding approach is to fire coal and natural gas. Natural gas costs are now closer to coal prices on a Btu basis with prices expected to remain low. Gas is easy to handle, available in large quantities, and has a higher H_2/C ratio than any fuel. Several developers have proposed dual coal-gas gasification, but no developer until now has offered a gasification technology that takes full advantage of the dual fuel firing to simplify both the gasifier and the full plant, to maximize flexibility and total efficiency, and to minimize the cost of product electricity (by IGCC) or liquids by LHS or other means.

GTI has developed the coal-natural gas fired hybrid molten bed (HMB) gasification technology to maximize IGCC power, liquid fuels by LHS, or chemicals yields with the highest possible overall plant efficiency and flexibility, lowest product cost, and least environmental impact. The HMB gasifier contains a bed of molten slag at 2500-2700°F maintained by fuel-rich combustion of natural gas and oxygen fired directly into the molten bath. Coal charged to the gasification chamber from above gasifies while circulating in the molten slag utilizing the excess heat from the partial oxidation of the fuel gas. The gasifier walls are built of tube banks with a thin layer of castable refractory on the inside. A thin layer of frozen slag forms on the walls, protecting them from abrasion so lifetime is indefinitely long, a process demonstrated with many mineral melts in GTI submerged combustion melters. Water passed through the wall tube banks forms useful steam, boosting overall efficiency. That steam is injected into the gasifier, providing additional reactant for steam carbon reactions while recuperating heat lost from the gasifier walls back to the gasifier. Syngas, steam from the walls, and fuel gas partial combustion products mix and exit the gasifier at 2500-2700°F.

HMB calculations and comparisons are promising compared with published DOE cases. For the IGCC case with MT subbituminous coal, oxygen demand has been calculated to be 8% lower, making the ASU smaller, and the H_2+CO yield is 15% greater, making electricity generation higher. The much lower HMB CO yield ($H_2/CO > 6$) leads to a 75% reduction in water gas shift (WGS) reactor size and WGS steam demand.

For the FT case with IL #6 coal, oxygen demand has been calculated to be 10% lower, making the ASU smaller, and the H_2+CO yield is 17% greater, making the FT diesel and naphtha yields higher. Producing HMB syngas with the optimum H_2/CO ratio of 2.1 leads to higher yields of liquids along with smaller WGS and auto-thermal reforming equipment and utility requirements inside the FT process.

In one novel configuration (see HMB/LHS version flow diagram in Figure 2) the combustion gases do not mix with the coal, and the coal carbon-steam reaction generates a higher quality syngas. Hot syngas provides heat to generate steam for injection and to endothermically reform a portion of the natural gas. The reformed gas is charged as fuel to the gasifier. This innovative method of recovering heat from hot syngas improves overall plant efficiency since there is no other need for this heat. This feature is possible only in a hybrid gasifier because coal-only gasifiers are not designed to accept fuel gas.

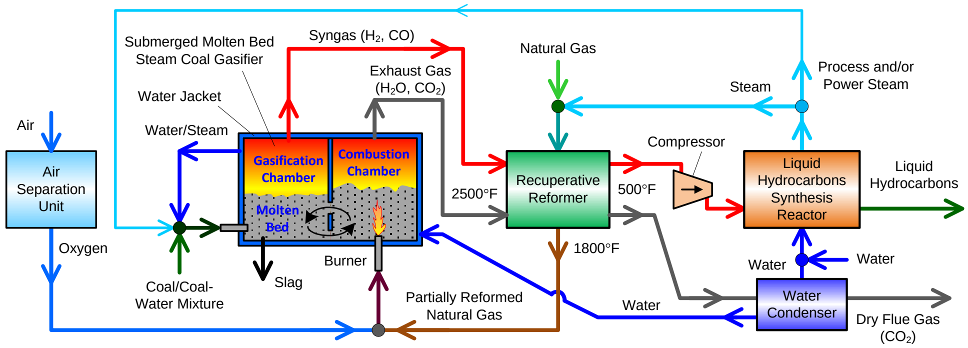


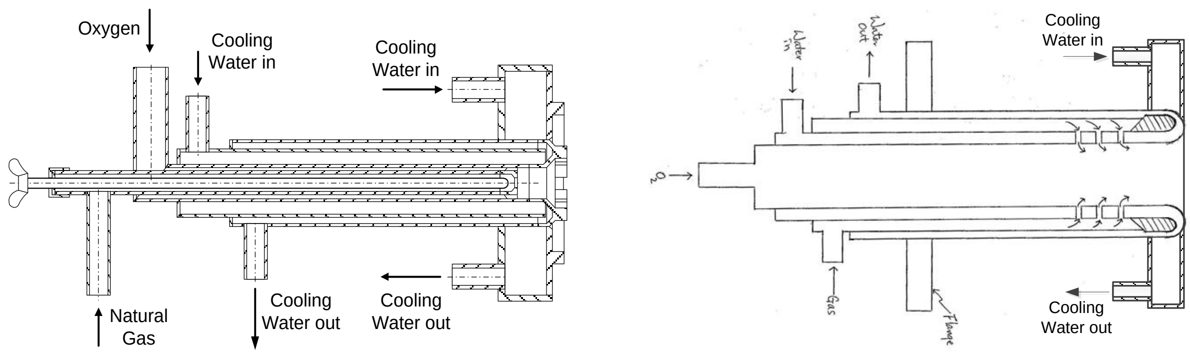
Figure 2. HMB/LHS Concept – One possible optimal configuration of HMB gasification

At the start of the proposed project, different HMB configurations were evaluated in order to optimize syngas quality and product and to lower overall product cost. One configuration recently evaluated considers an HMB gasifier in which combustion product gases do not mix with the gasification products and gasification is driven entirely by the steam-carbon reaction. While potentially more physically complex, this configuration has been calculated to need only 27% natural gas (73% coal) and a 2:1 steam to carbon ratio to generate a syngas with a H2/Co value of 2.0 and a yield of gasoline plus LPG of 54%. These promising results confirm that an optimum HMB configuration can significantly improve liquid yield from a gasification process. Promising HMB configurations were analyzed during the design of the laboratory HMB gasifier facility.

Experimental Methods

The work in this project involved techno-economic analyses and laboratory testing of the HMB gasification concept. Experimental work was conducted to study the production of syngas when firing fuel gas (natural gas or syngas with or without steam addition) under substoichiometric conditions into a bed of molten slag while simultaneously charging pulverized coal to the top of the molten bed of slag. Design and setup of that equipment was part of this project. Much of the needed equipment was already available at GTI and was repurposed for use in this study. The HMB gasifier test unit and some support components were specially designed and built for this project. All materials were purchased under the DOE funds to this project.

GTI previously developed, designed, fabricated, tested, and successfully commercialized patented oxygen-natural gas burners (see Figure 3) for rising into a bed of molten material in the submerged combustion melter (SCM). The SCM technology has been used to melt a wide range of mineral materials including mineral wool, cement kiln dust, electric arc furnace dust, simulated high level radioactive waste, and a wide range of industrial glasses (fiberglass, container glass, etc.). The HMB process uses the burner to fire natural gas or syngas with steam into the bed of molten slag under substoichiometric conditions. Firing rate of the burners is 0.5-1 MMBtu/h which is appropriate for laboratory demonstration of HMB gasification in a molten bed. Coal feed rates were set at 20-40 pounds per hour to the top of the molten bed in the simulated gasifier.



(a) *(b)*

Figure 3. GTI's Water-Cooled Burners for Oxy-Fired Submerged Combustion Melter: (*a*) Burner With Center Nozzle for Natural Gas; (*b*) Burner With Peripheral Nozzles for Natural Gas

Testing was carried out in the GTI Combustion Laboratories and used much of the equipment designed and built for testing submerged combustion glass melting. The oxygen and gas supplies, the exhaust duct, the baghouse, the sensors and controls, and other equipment was used in order to maximize the work that could be accomplished with the existing HMB gasification project budget. The layout of the HMB test gasification unit is shown in Figure 4. The photographs in Figure 5 show the burner, the HMB test chamber, the blending station to generate syngas, and the coal feeder.

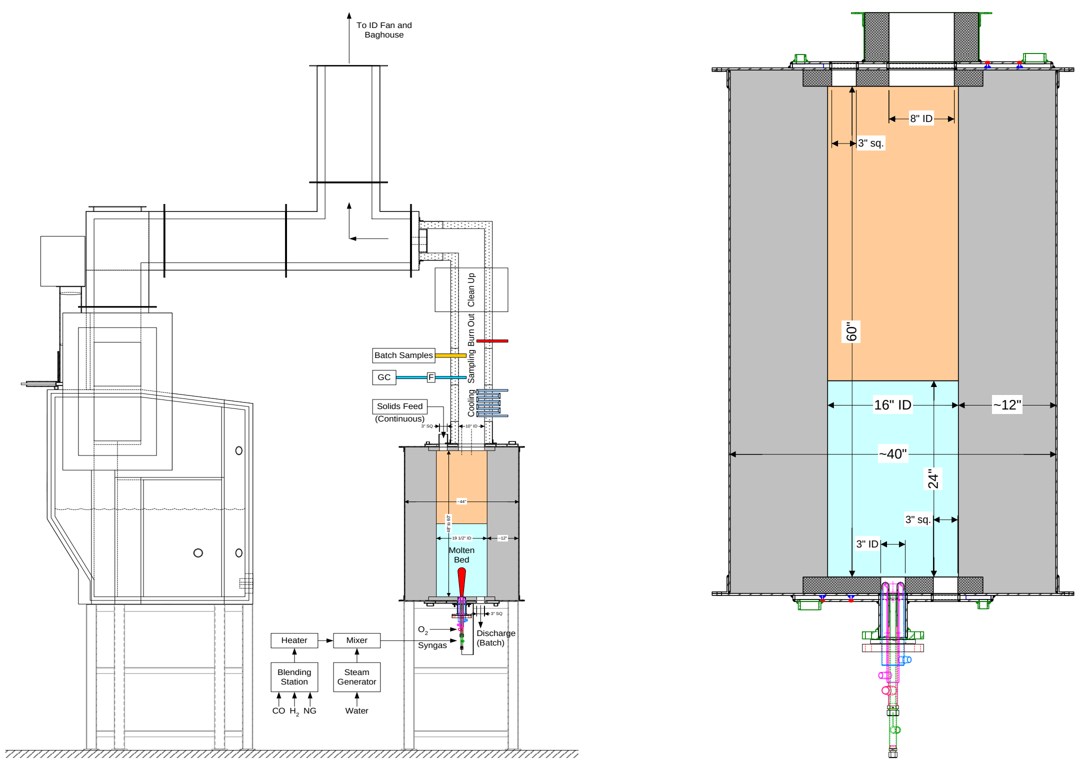


Figure 4. Laboratory HMB Test Unit

Figure 5. Photographs of the Oxy-Fuel Gas Burner, the HMB Gasification Test Unit, the Fuel Gas Blending Station, and the Coal Feeder

The testing equipment was scaled and sized. The team used a single oxy-gas burner firing upward into a bed of 375 pounds of molten slag. Coal feed rates were 20 to 40 pounds per hour. Gas analysis equipment, including, a dedicated gas chromatograph was set up for testing. The gasifier tube had a circular cross section and was lined with 9 inches of cast refractory so the high temperatures of the molten bed could be established and maintained. The final hook-up of components and shakedown testing was carried out this quarter before testing was conducted. A source of pulverized (70 micron) Illinois #6 coal was identified, and the coal needed for project testing was stored in sealed drums in preparation for testing in the project.

Testing was delayed while a determination was made regarding the possible need for approval from EPA to allow laboratory testing of a bench-scale unit. The approval process was completed with the determination that this testing fell under the umbrella of laboratory-scale testing needing no special permit. This allowed testing to be conducted in the final project quarter. A six-month no-cost time extension was requested and granted by DOE NETL to allow for this environmental permit process to be completed.

Testing began with charging 375 pounds of glass cullet to the test chamber. The burner, operating on natural gas and oxygen, created a molten bed of glass. Once steady conditions were established, independent variables of natural gas rate, oxygen rate, and coal rate were varied. In later tests, steam was added to the natural gas, while other tests changed the natural gas to two syngas compositions simulating reformed syngas.

The photographs in Figure 6 show the top of the test chamber during testing and the test chamber melt discharge of molten slag after a test. Oxygen was injected above the bed and also above the gas chromatograph sampling port. This created the combustion zone clearly visible in the photograph. This oxygen was used to burn out all CO and H2 before the product gas was vented to atmosphere. The molten slag was discharged at the end of a day of testing. The slag was collected in hoppers containing sand to handle the high heating of the cooling slag. The slag was removed through a port in the bottom of the gasification test unit. A brick wedged into the port was carefully removed, and the melt discharged as a steady stream through the open port. No extra effort was needed to keep the port clear during discharge. This procedure led to complete evacuation of the melt chamber. The hot refractory walls kept all slag fully molten until the slag was discharged from the gasification test unit.

Figure 6. Photographs of the Top of the HMB Gasification Unit During Testing Showing Gas Burn-out and the Removal of Molten Slag After Completing Testing

The first HMB test conditions were carried out with oxygen-natural gas and coal. No steam was injected through the burner. This enabled engineers to obtain a good understanding of equipment operation and to establish a gasification baseline. The second series of tests were made with natural gas and steam. The third test series was made with two different syngas compositions and steam. The syngas compositions were selected to approximate the composition of HMB product syngas after partial reforming. Operating conditions included:

- One oxygen-natural gas burner – 300 SCFH
- O2/NG ratio – 2.1-1.6 of stoichiometric ratio
- Inlet energy content – 49-57% from coal
- Fuel gas – house natural gas and two syngas mixtures containing CO, CO2, H2, H2O, and CH4
- Bed – 375 lb molten glass at ~2450ºF initially with coal slag added during testing
- Coal – pulverized (200 mesh) Illinois #6 at 24-36 lb/h
- Coal transport by nitrogen carrier gas
- Syngas analysis by on-line gas chromatograph (H2, CO, CO2, CH4, N2, O2)

Results and Discussions

Nexant, Inc. was a project partner for conducting HMB process techno-economic analyses (TEA). Two complete TEA analyses were conducted, one looking at HMB gasification for IGCC power production and one for HMB for Fischer-Tropsch (FT) production of diesel. Both gasification plant configurations assumed capture of carbon dioxide for later sequestration. The Shell entrained flow gasifier was used for comparison in both the IGCC and FT configurations. The Shell gasification process is a good baseline for comparison because temperatures are similar to HMB, and both processes operate in a slagging regime. Detailed analyses of the Shell gasification process have been carried out for NETL and that resulting report provided cases that were used for comparison in this project. The Shell IGCC configuration was recalculated by Nexant engineers. Good agreement was found between the earlier TEA reported information and the Nexant calculated results. A similar NETL report for Shell gasification as part of a FT plant was not finished in time for use as a background TEA for comparison. For the FT configuration Nexant engineers calculated the Shell configuration and used those results as the baseline case.

The two TEA configurations, including sensitivity analyses, were prepared as stand-alone reports by Nexant. These reports are attached to the report as Appendices. Shown below are summaries of the results of the two TEA analyses.

The techno-economic analysis follow the guideline and procedures in the following DOE/NETL Reports:

1. "Cost and Performance for Fossil Energy Plants, DOE/NETL-2010/1397 Volume 1: Bituminous Coal and Natural Gas to Electricity" for design basis preparation and economic evaluation methodology as needed.
2. Supplement to Volume 1, "Updated Costs (2011 Basis), DOE/NETL
3. Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis, October 2014, DOE/NETL-2010/1396 was not available at the time the analyses were performed, so the 2007 DOE report titled "Baseline Technical and Economic Assessment of a Commercial Scale FT Liquid Facility" was used instead
4. NETL's Series of Quality Guidelines for Energy System Studies (QGESS)

IGCC and FT plant overall heat and material balances are being carried out using a Nexant in-house spreadsheet model with verification/benchmark of selected process units by Aspen, Hysys, or GateCycle simulation models. Equipment sizing and costs, and final integrated overall performance and cost were completed using a Nexant in-house spreadsheet model

Nexant needed to repeat the DOE IGCC case S1B with a Shell gasifier to confirm the basis for the HMB IGCC calculations. In-house modeling of the overall process was performed, including gas turbine and steam turbine performance, using:

- DOE/NETL 2010/1399 S1B as case reference
- "QGESS Process Modeling Design Parameters" for design bases assumptions

Overall heat and material balance (H&MB) and overall utility balances defined the process plant capacities and balance of plants (BOP) for estimating auxiliary power consumptions via capacity prorating from the DOE S1B case. All costs are listed in 2011 dollars.

Work on the IGCC cases began with carrying out the baseline Shell gasification case calculations. Initial conditions for the DOE S1B and repeat Nexant case 1a are given below.

- DOE S1B case:
  - Costs were escalated from 2007 to 2011 dollars using escalation factors obtained from "Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases" report (DOE/NETL-341/082312)
  - ~20% cost increase from 2007 to 2011
- Nexant 1a case:
  - CAPEX was estimated using escalated S1B 2011 costs as reference basis
  - Established relevant scaling exponents and reference parameters as prescribed by DOE QGESS "Capital Cost Scaling Methodology" document
  - Scaled capital costs for process plants and BOP systems based on capacities defined from overall H&MB and utility balances

Nexant has evaluated five IGCC gasification configurations for IGCC Power option and the FT Transportation Fuel option.

1. Gasifier Only - 100% coal feed to the gasifier (reference model with Shell gasifier); No natural gas feed or steam; No Reformer
2. Gasifier Only - Co-feeds of coal, natural gas and steam to the gasifier); No reformer; Recycled gas quenched syngas
3. Gasifier with Pre-Reformer - Coal feed to the gasifier; Natural gas and steam feed to the syngas heated reformer; reformate to the gasifier
4. Gasifier with Aft-Reformer - Coal feed to the gasifier; Natural gas, steam and gasifier syngas to the reformer; Reformate to heat recovery
5. Gasifier with Parallel Reformer - Coal feed to the gasifier; Natural gas plus steam to the syngas heated reformer

Reference cases for these evaluations are as follows:

- IGCC Power Option – Cost and Performance Baseline for Fossil Energy Plants - Volume 3a: Low Rank Coal to Electricity: IGCC Cases, May 2011, DOE/NETL-2010/1399 – Case S1B
- FT Transportation Fuel Option – Cost and Performance Baseline for Fossil Energy Plants - Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis, October 2014, DOE/NETL-2010/1396 – Note that this document was not available when this report was first drafted

The evaluation has concluded that configuration 1, 2 and 3 will meet the technical requirements of the IGCC option and configurations 2 and 3 will meet the technical requirements for the FT options. These configurations also potentially have the most competitive COE and COPs Configurations 4 and 5 are not considered for further COE or COP evaluations as these cases do not meet the selection criteria on technical or cost basis. A discussion of the technical requirements and the selection criteria are included in the attached report.

Nexant evaluated five Molten Bed Gasification (MBG) configurations and their application to IGCC power production and Fischer-Tropsch (FT) transportation fuels options.
The five configurations are as follows:

1. Gasifier Only - 100% coal feed (Shell IGCC reference case)
    a. No natural gas feed or steam
    b. No steam reformer
2. Gasifier Only - coal and natural gas feeds (GTI configuration)
    a. Natural gas and steam are co-fed into the gasifier
    b. No steam reformer
3. Gasifier with Pre-reformer
    a. Coal feed to the gasifier
    b. Natural gas plus steam to the steam pre-reformer
    c. Reformate to the gasifier
    d. Reforming duty is provided by gasifier syngas at a temperature approach of 50°F to the reforming temperature
4. Gasifier with Aft-reformer
    a. Coal feed to the gasifier
    b. Natural gas, steam and gasifier syngas to the steam aft-reformer
    c. Reforming temperature is controlled by coal/NG feed ratio
    d. Requires sulfur tolerant reforming catalyst
5. Gasifier with Parallel reformer
    a. Coal feed to the gasifier
    b. Natural gas plus steam to the steam parallel reformer
    c. Reforming duty is provided by gasifier syngas at a temperature approach of 50°F to the reforming temperature

These gasification configurations were modeled based on a range of steam/carbon ratios and coal to natural gas feed mix ratio. The results are evaluated based on the following key criteria for the IGCC and the FT options:

1. IGCC Power Production
    a. High cold gas efficiency
       Cold gas efficiency is defined as $(H_2+CO)_{HHV}/(\text{Coal + Natural Gas})_{HHV}$
    b. Low oxygen consumption
    c. Low steam consumption
    d. Low CH_4 in fuel gas to turbine
2. FT Transportation Fuels
    a. Meets FT feed specification without WGS
        i. $H_2/CO = 1.5$ for iron based FT catalyst
        ii. $H_2/CO = 2$ for cobalt based FT catalyst
        iii. $\text{Inert}/(H_2+CO) < 0.1$ (Inert includes CH_4, N_2, & Ar)
    b. High cold gas efficiency
    c. Low oxygen consumption

Techno-Economic Analysis – IGCC Configuration

Each configuration is modeled based on producing a syngas with the same HHV as the Shell IGCC reference case. Nitrogen conveying is used because of its availability from the ASU. For each configuration, an optimum case is selected for comparison with other configurations. A summary of all the configurations is shown in the table below.

Configurations 1, 2 and 3 met the selection criteria with configuration 3 being the best IGCC configuration for the following reasons:

- Highest cold gas efficiency of 89.9%
- Lowest oxygen demand at 3,008 TPD
- Steam demand is the lowest for the gasifier/reformer configurations (319,000 lb/h)

Configuration 4 has the lowest cold gas efficiency due to the high CH_4 content in the Aft-reformer syngas and it also requires a sulfur tolerant reforming catalyst which we believe is not yet commercially available. Configuration 5 requires a minimum of 85% coal/15% NG feed mix to provide adequate reformer duty. These two configurations are not good fits for the IGCC option selection criteria. The IGCC techno-economic analysis examined the Shell base case and HMB cases 1, 2, and 3 since those were previously determined to be the most promising HMB cases. Overall block diagrams for the four cases are shown in Figures 7-10. A summary of the cases considered is shown in Table 1.

The overall block flow diagram (BFD) for the Shell slagging SGCP gasifier IGCC case is shown in Figure 7.

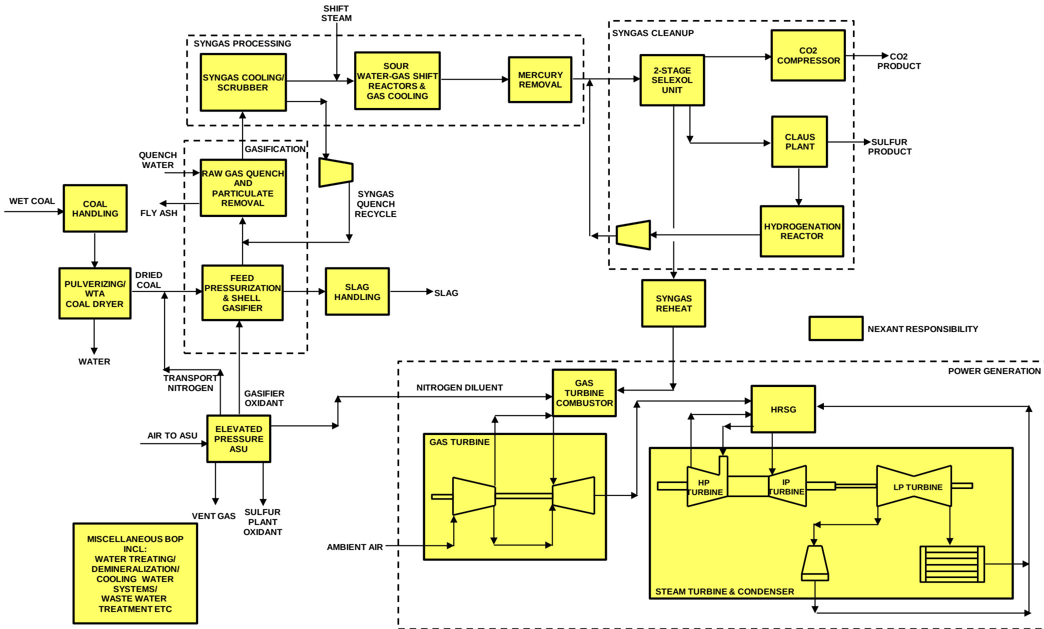


Figure 7. BFD of the Shell Slagging Gasifier in IGCC Mode

The overall BFD for HMB Case 1 for IGCC with CO2 capture and using 100% PRB coal is shown in Figure 8.

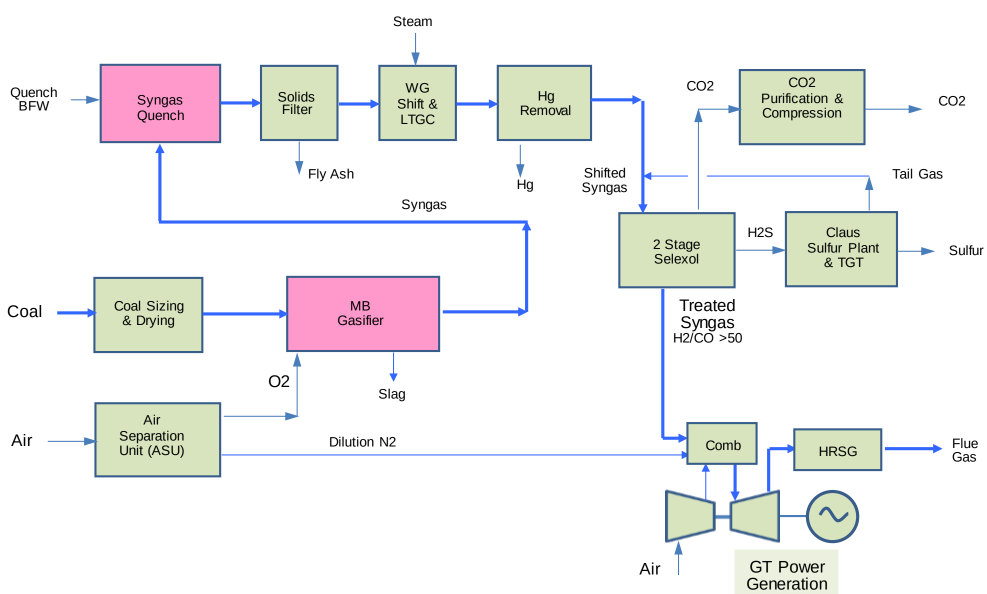


Figure 8. BFD of the HMB Gasifier in IGCC Mode With CO2 Capture – Case 1

The overall BFD for HMB Case 2 for IGCC with CO2 capture and using 55% PRB coal and 45% natural gas is shown in Figure 9.

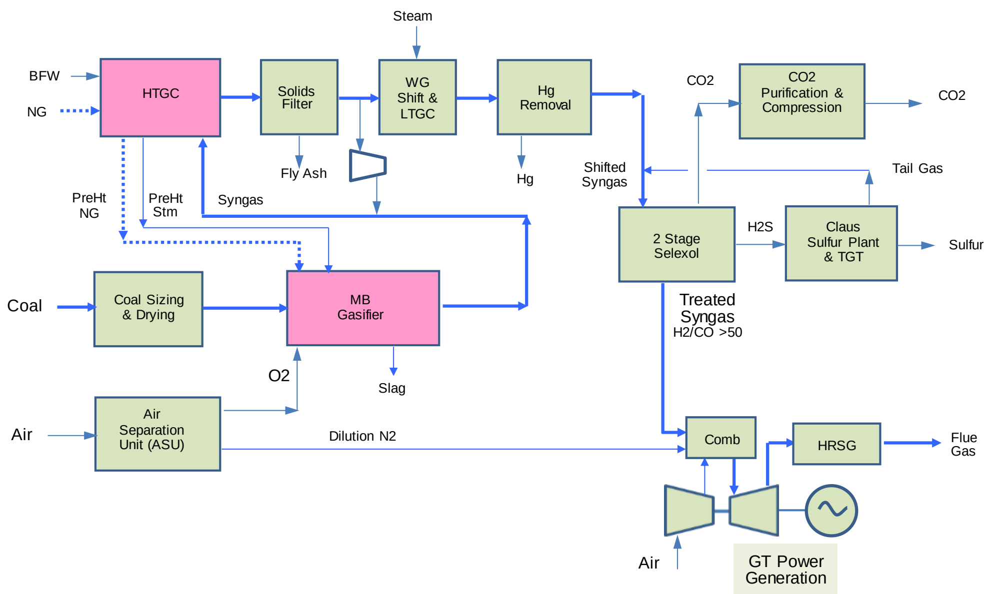


Figure 9. BFD of the HMB Gasifier in IGCC Mode With CO2 Capture – Case 2

The overall BFD for HMB Case 3 for IGCC with pre-reformer, CO2 capture, and using 55% PRB coal and 45% natural gas is shown in Figure 10.

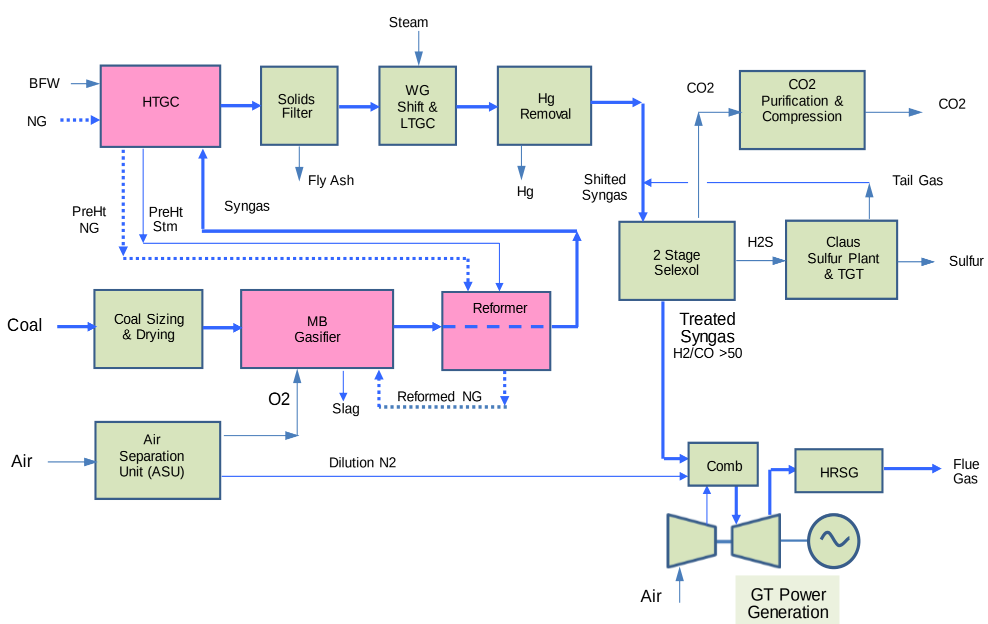


Figure 10. BFD of the HMB Gasifier in IGCC Mode With CO2 Capture – Case 3

Table 1. IGCC Techno-Economic Analysis Basis

	Reference Case[1]	Case 1	Case 2	Case 3
Gasification Technology				
Shell Dry Feed Gasification (SCGP)	X			
GTI Molten Bed Gasifier (MBG)		X	X	Hybrid MBG (with Pre-Reformer)
Feed Mix	100% Coal	100% Coal	55% Coal / 45% NG	55% Coal / 45% NG
Coal Type	PRB	PRB	PRB	PRB
Steam to Gasifier			X	
Syngas Cooling				
Recycle gas quench	X		X	
Water quench	X	X		
Steam Generation	X	X	X	X
Feed Preheat			X	X
Reformer Reaction Heat				X
Syngas Cleanup				
AGR - Selexol[2]	X	X	X	X
Water Gas Shift				
High & Low Temperature WGS	X	X	X	X
Gas Turbine				
Advanced Turbines	X	X	X	X
CO_2 Purification and Compression	X	X	X	X

1 Nexant's Simualtion of the DOE S1B reference case.

2 Additional trace contaminant cleanup such as mercury removal will be included as defined by DOE/NETL baseline studies.

The IGCC techno-economic analysis summary is shown in the three tables below. Table 2 summarizes overall plant characteristics. Table 3 provides the power summary and overall plant efficiency. Table 4 provides the cost of electricity (COE) summary results.

Table 2. TEA Summary for IGCC Configuration – Overall Plant Characteristics

Case	Reference Case	Case 1	Case 2	Case 3
Source	Nexant Modeling	Nexant Modeling	Nexant Modeling	Nexant Modeling
Gasifier & Coal Feed Technology	Shell SCGP Gasifier	GTI MBG	GTI MBG	GTI Hybrid MBG
Coal Type	PRB	PRB	PRB	PRB
Feed Mix, % HHV	100% Coal	100% Coal	55% Coal / 45% NG	55% Coal / 45% NG
As-Received Coal Feed, lb/hr	585,971	580,414	334,168	300,991
Natural Gas Feed, lb/hr	-	-	103,680	93,386
Carbon Capture, %	90	90	90	90
Cold Gas Efficiency, %	80.6	81.6	78.0	87.4
Acid Gas Recovery Technology	Selexol	Selexol	Selexol	Selexol

Table 3. TEA Summary for IGCC Configuration – Power Summary and Overall Plant Efficiency

Case	Reference Case	Case 1	Case 2	Case 3
Power Summary, MWe				
Power Generation :				
Gas Turbine	*430.0*	*432.1*	*431.3*	*430.0*
Steam Turbine	*222.2*	*211.1*	*267.6*	*207.4*
Total Gross Power	*652.2*	*643.2*	*698.9*	*637.4*
Auxiliary Load Total	*192.6*	*188.7*	*189.0*	*154.1*
Net Power Generation	459.6	454.5	509.9	483.3
Net Plant Efficiency, % HHV	31.2%	31.2%	33.4%	35.2%

Table 4. TEA Summary for IGCC Configuration – Cost of Electricity

Case	Reference Case	Case 1	Case 2	Case 3
Capacity Factor (CF), %	80	80	80	80
Net Power Generation, MWe	459.6	454.5	509.9	483.3
2011 Capital Cost, $MM				
Total Plant Cost, $MM	2,017	1,643	1,584.2	1539.0
Total Overnight Cost, $MM	2,472	2,020	1,938	1882.3
2011 Operating Cost, $MM/yr				
Fixed Operating Costs	*74.0*	*61.8*	*59.9*	*58.4*
Variable Operating Costs @ 100% CF	*56.4*	*48.2*	*43.8*	*41.9*
Fuel Costs @ 100% CF, Coal @$19.63/ton	*50.4*	*49.9*	*28.7*	*25.9*
NG @ $5/MMBtu	*0*	*0*	*102.0*	*91.9*
Cost of Electricity (excl TS&M), mills/kWh	144.8	122.9	123.2	124.1
Cost of Electricity (incl TS&M), mills/kWh	161.6	139.6	135.7	135.9

The summary results show that case 1 has the same efficiency as the baseline case but a reduction in electricity cost, primarily through lower capital costs. Cases 2 and 3 show an increases in plant overall efficiency and even larger decreases in COE. While the pre-reformer in case 3 provides higher efficiency, the COE is the same as case 2. The use of natural gas in cases 2 and 3 raises fuel costs. But the higher hydrogen content leads to smaller water gas shift demand and better steam management. Overall, natural gas also allows for lower total fuel demand and a decrease in CO2 production. Table 5 summarizes the plant performance. Tables 6-10 provide data from the calculations for key plant systems.

Table 5. IGCC Plant System Costs

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Reference Case	Case 1	Case 2	Case 3
Gas Turbine Power	429,974	432,063	431,306	430,022
Steam Turbine Power	222,181	211,142	267,585	207,376
TOTAL POWER, kWe	**652,155**	**643,205**	**698,890**	**637,397**
AUXILIARY LOAD SUMMARY, kWe				
Coal Handling	510	505	291	262
Coal Milling	2,730	2,704	1,557	1,402
Slag Handling	580	483	278	250
WTA Coal Dryer Compressor	9,370	9,281	5,344	4,813
WTA Coal Dryer Auxiliaries	620	614	354	318
Natural Gas Compressors			5,658	5,078
Gasifier Steam Generator Circ. Pumps		196	195	195
Air Separation Unit Auxiliaries	1,003	974	1,051	777
Air Separation Unit Main Air Compressor	63,719	61,908	66,805	49,348
Oxygen Compressor	8,830	9,336	10,278	7,547
Nitrogen Compressors	33,340	31,572	32,463	28,026
CO_2 Compressor	31,544	31,173	26,159	23,493
Boiler Feedwater Pumps	3,851	3,593	4,886	4,115
Condensate Pump	194	247	262	219
Quench Water Pump	760	760	0	0
Syngas Recycle Compressor	820	0	1,116	0
Circulating Water Pump	2,931	2,849	3,366	2,771
Ground Water Pumps	320	371	341	299
Cooling Tower Fans	1,911	1,858	2,195	1,807
Air Cooled Condenser Fans	2,771	2,505	3,268	2,495
Scrubber Pumps	20	20	17	15
Acid Gas Removal	18,390	18,199	15,274	13,740
Gas Turbine Auxiliaries	998	1,003	1,001	998
Steam Turbine Auxiliaries	96	91	115	89
Claus Plant/TGTU Auxiliaries	249	247	142	128
Claus Plant TG Recycle Compressor	1,517	2,770	2,153	1,964
Miscellaneous Balance of Plant	3,000	2,972	1,711	1,541
Transformer Losses	2,507	2,472	2,686	2,450
TOTAL AUXILIARIES, kWe	**192,581**	**188,704**	**188,964**	**154,142**
NET POWER, kWe	**459,574**	**454,501**	**509,926**	**483,255**
Net Plant Efficiency, % (HHV)	31.2%	31.2%	33.4%	35.2%
Net Plant Heat Rate, Btu/kWh	10,919	10,934	10,205	9,699
CONDENSER COOLING DUTY, MMBtu/hr	**1,170**	**1,058**	**1,380**	**1,053**
CONSUMABLES				
As-Received Coal Feed, lb/hr	585,971	580,414	334,168	300,991
Thermal Input, kWt	1,470,705	1,456,405	1,525,059	1,373,649
Raw Water Withdrawal, gpm	3,520	4,744	4,270	3,772
Raw Water Consumption, gpm	2,842	4,074	3,740	3,286

Table 6. IGCC Plant Fuel Requirements and Efficiencies

	Reference Case	Case 1	Case 2	Case 3
Feed HHV, MMBtu/hr	5,019	4,971	5,205	4,688
Δ% of Reference Case	-	-1%	+3.7%	-6.6%
Net Plant Heat Rate, Btu/kWh	10,919	10,934	10,205	9,699
Cold Gas Efficiency, %	80.6	81.6	78.0	87.4
Net Plant Efficiency, % HHV	31.2%	31.2%	33.4%	35.2%

IGCC plant performance results showing the overall fuel requirement and efficiency is shown in Table 6. The basis is a nominal 430 MWe gas turbine (with the total for 2 trains). In cases 2 and 3 there is a significant decrease in net plant heat rate (Btu/kwh).

The air separation unit (ASU) demand is shown in Table 7. ASU capacities are for a total of two trains. Due to the smaller ASU capacity for case 3 with the need to maintain 430 MWe gas turbine output, an increase in syngas H_2+CO rate (1% higher) is required to compensate for the reduction in dilution nitrogen. Only case 3 provides a large decrease in ASU capacity.

Table 7. IGCC Case Air Separation Unit Demand

	Reference Case	Case 1	Case 2	Case 3
95% O_2, Tons/D	4,335	4,223	4,557	3,366
Δ% of Reference Case	-	-2.6%	+5.1%	-22.3%
Dilution N2 to GT, Tons/D [1]	12,689	12,341	12,689	10,955
Δ% of Reference Case	-	-2.7%	-	-13.7%

The water gas shift (WGS) results using low temperature and high temperature shift are shown in Table 8. Case 1, coal only, has the same syngas H2/CO as the baseline case, but the value is much higher for the more hydrogen-rich cases 2 and 3.

Table 8. Water Gas Shift Results for IGCC Cases

	Reference Case	**Case 1**	**Case 2**	**Case 3**
Inlet Temperature, °F	449	450	450	450
Injected Steam, lbs/hr	175,564	175,630	200,218	249,473
Outlet H2/CO, mol/mol	56	53	66	67

The CO2 capture results are shown in Table 9. Selexol AGR is used in all cases. The AGR plant size is a function of the feed syngas flow and the CO_2 partial pressure. Lower CO_2 partial pressures require higher absorbent flow. The lower syngas flows for cases 2 and 3 are offset by the higher absorbent flows for these cases due to the lower CO_2 partial pressures. For the same GT fuel gas heating value requirement, less CO_2 is produced for gasifier feed containing natural gas in the feed than for the 100% coal feed. Hence, less CO_2 compression capacity is required for cases 2 and 3.

Table 9. CO2 Capture Results for IGCC Cases

	Reference Case	Case 1	Case 2	Case 3
CO_2 Captured,%	90%	90%	90%	90%
Capacity, MMSCFD Syngas	551	541	499	482
ppCO_2, psia	195	192	175	163

	Reference Case	Case 1	Case 2	Case 3
CO_2 Capacity, Tons/D	11,561	11,420	9,583	8,606
Δ% of Reference Case	-	-1.2%	-17.1%	-25.6%
CO_2 Compression , kWe	31,544	31,173	26,159	23,493

The gas turbine (GT) results are shown in Table 10. Less dilution nitrogen is available in case 3 due to the smaller ASU size. Fuel gas rate is increased by 1% to maintain the same 430 MWe GT output.

Table 10. Gas Turbine Results for IGCC Cases

	Reference Case	Case 1	Case 2	Case 3
Fuel Gas LHV, Btu/SCF	236	238	245	246
Fuel Gas LHV After Dilution, Btu/SCF	118	121	121	130
Dilution N2, Tons/D	12,689	12,341	12,689	10,955[1]
GT Exhaust Temperature, °F	1,042	1,055	1,054	1,084
GT Generator Output, MWe	430	432	431	430

The total plant cost summary for the baseline case and cases 1 through 3 are shown in Table 11. Cost savings are realized from the smaller HMB gasifier in cases 1 through 3 and from less coal handling and prep in cases 2 and 3.

The operating cost summary for all cases is shown in the table in Table 12. The same trend is reflected in smaller HMB gasifiers but higher costs for fuel in the hybrid cases 2 and 3. Overall, the HMB process has lower operating costs.

Sensitivity analyses were carried out for coal price, natural gas price, CO2 sale price, and cost of CO2 emissions. Figure 11 shows that when coal price varies and all other costs are constant, the COE for the HMB cases remains lower the Shell COE at all times.

Table 11. Total Plant Summary for IGCC Cases

CODE OF ACCOUNTS TOTAL PLANT COST, 2011 $MM	Reference Case	Case 1	Case 2	Case 3
1. COAL & SORBENT HANDLING	49.4	49.1	34.8	32.7
2. COAL & SORBENT PREP & FEED	237.8	236.3	164.1	153.2
3. FEEDWATER & MISC BOP SYSTEMS	34.4	29.9	36.1	29.3
4. GASIFIER & ACCESSORIES	751.4	401.5	432.2	453.4
5A. GAS CLEANUP & PIPING	289.9	289.4	269.7	268.8
5B. CO2 REMOVAL & COMPRESSION	66.3	65.6	56.2	51.1
6. COMBUSTION TURBINE/ACCESSORIES	159.4	159.4	159.4	159.4
7. HRSG, DUCTING & STACK	54.0	53.5	54.2	54.3
8. STEAM TURBINE GENERATOR	122.5	112.1	140.6	114.1
9. COOLING WATER SYSTEM	27.0	27.9	28.3	25.2
10. ASH/SPENT SORBENT HANDLING	44.4	39.5	28.1	26.3
11. ACCESSORY ELECTRIC PLANT	105.0	104.1	105.7	97.6
12. INSTRUMENTATION & CONTROL	32.0	31.9	31.9	31.1
13. IMPROVEMENTS TO SITE	22.5	22.1	22.1	22.0
14. BUILDINGS & STRUCTURES	20.9	20.9	20.8	20.5
TOTAL TPC	2,016.6	1643.2	1584.2	1539.0

Table 12. Operating Cost Summary for all IGCC Cases

OPERATING COSTS, 2011 $MM/yr	Reference Case	Case 1	Case 2	Case 3
FIXED OPERATING COSTS				
Annual Operating Labor Cost	7.2	7.2	7.2	7.2
Maintenance Labor Cost	19.5	15.9	15.4	14.9
Administration & Support Labor	6.7	5.8	5.6	5.5
Property Taxes and Insurance	40.3	32.9	31.7	30.8
TOTAL FIXED OPERATING COSTS	73.8	61.8	59.9	58.4
VARIABLE OPERATING COSTS (@ 100% CF)				
NON-FUEL VARIABLE OPERATING COSTS				
Maintenance Material Cost	45.3	36.9	35.6	34.6
Water	1.5	2.1	1.9	1.7
Chemicals				
MU & WT Chemicals	1.5	2.0	1.8	1.6
Carbon (Hg Removal)	0.1	0.1	0.1	0.1
WGS Shift Catalyst	1.2	1.2	0.7	0.7
Reformer Catalyst	0	0	0	Note 1
Selexol Solution	1.2	1.2	1.0	0.9
Claus Catalyst	0.1	0.1	0.1	0.1
Waste Disposal				
Spent Mercury Catalyst	0.0	0	0.0	0.0
Slag	5.4	4.5	2.6	2.3
FUEL (@ 100% CF)	50.4	49.9	130.8	117.8
TOTAL VARIABLE OPERATING COSTS	106.7	98.0	174.6	159.8
Note 1) Typical catalyst replacement frequency is 2 to 3 years				

Figure 12 shows the effect of changing natural gas cost on COE. Only HMB cases 2 and 3 have natural gas as an input. Comparison on COE with the figure above shows that HMB cases at the baseline coal cost have lower COE when natural gas cost increases to as much as $9/MMBtu. This confirms that the HMB gasification is promising in a wide range of economic conditions, even when natural gas costs rise well above their current, historically low values.

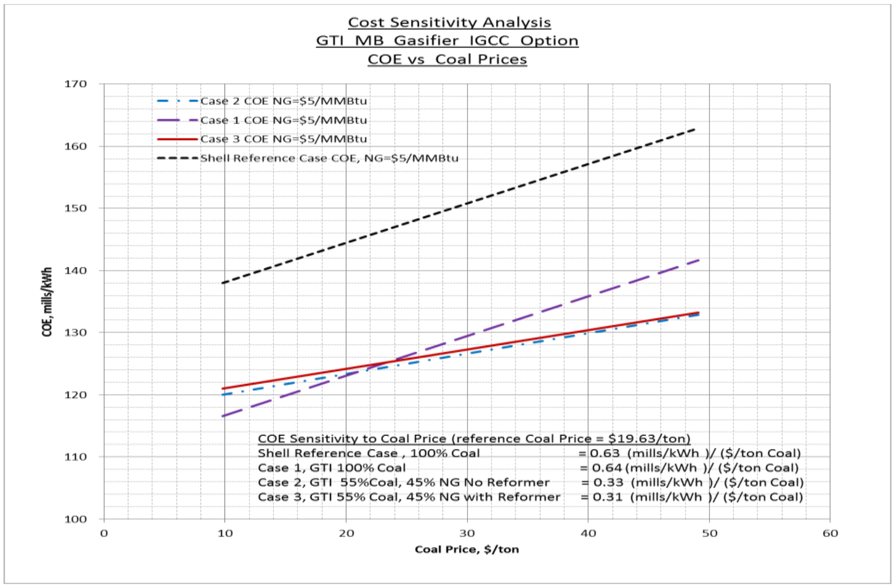


Figure 11. Cost Sensitivity Analysis – COE vs. Coal Price

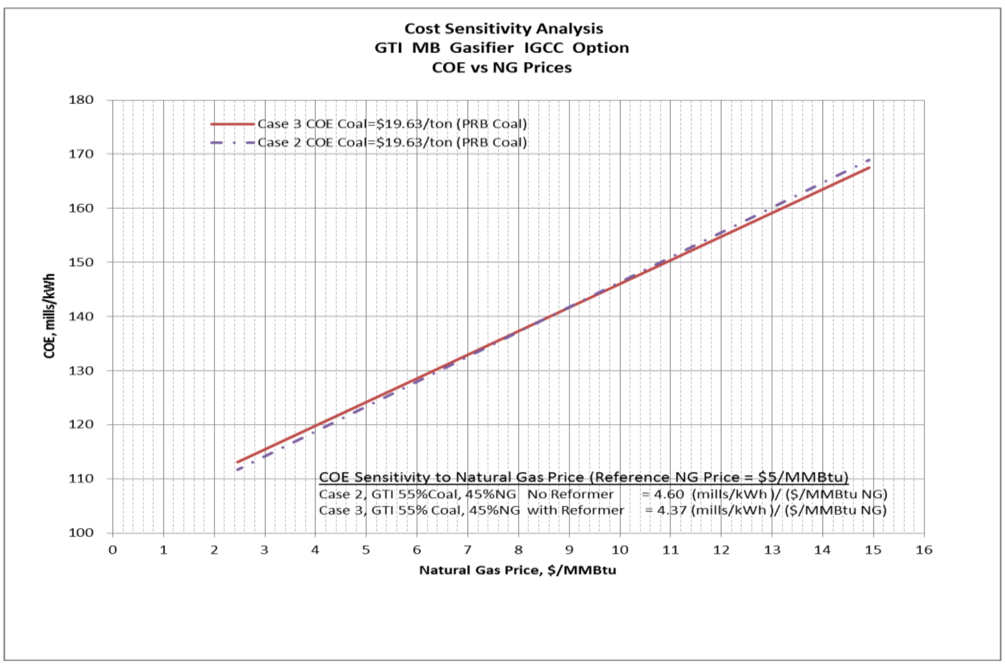


Figure 12. Cost Sensitivity Analysis – COE vs. Natural Gas Price

The initial assumption for all cases was a value for CO2 sales of $0/ton. However, if a market does exist for CO2, the sale could lower IGCC plant COE. Figure 13 shows that the Shell baseline case and HMB case 1 with 100% coal input have the same decrease in COE with increasing CO2 price. Cases 2 and 3 with coal-natural gas fuel produce less CO2, so benefits of CO2 sale are smaller. If CO2 price exceeds $40/ton, higher sales of CO2 will cause cases 1 and 2 to have lower COE than cases 2 and 3 with coal-natural gas fuel.

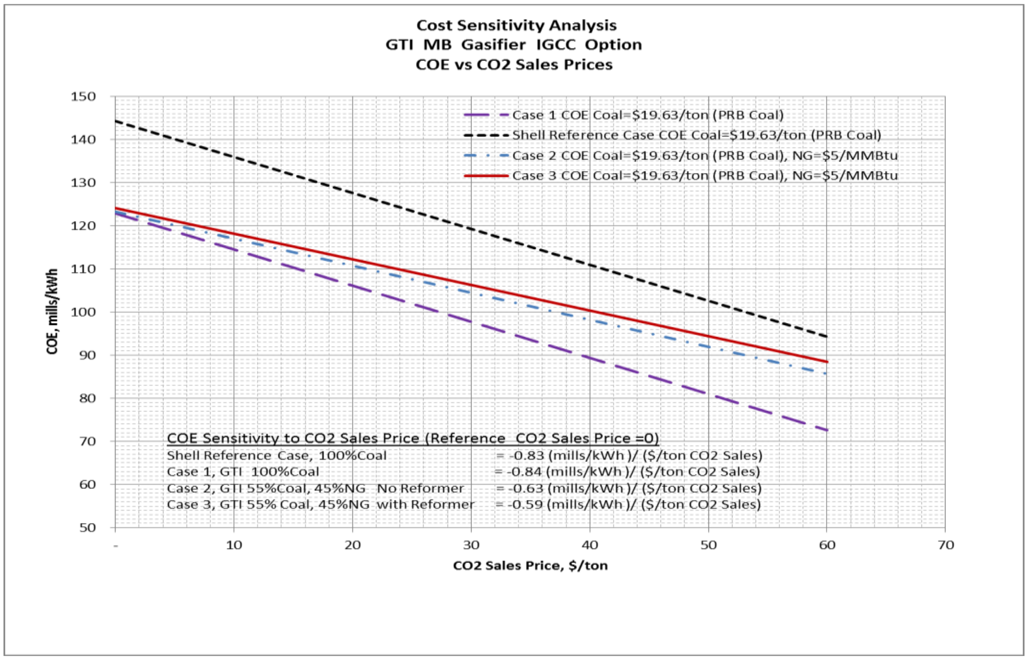


Figure 13. COE as a Function of CO2 Price for Shell and Case 1 HMB IGCC

Figure 14 considers the sensitivity of COE to the cost of CO2 emissions. Results show that the Shell IGCC case and the three HMB cases have similar sensitivity to CO2 emissions costs. Therefore, in all CO2 emission price situations, the HMB cases retain their COE advantages over the baseline Shell IGCC case.

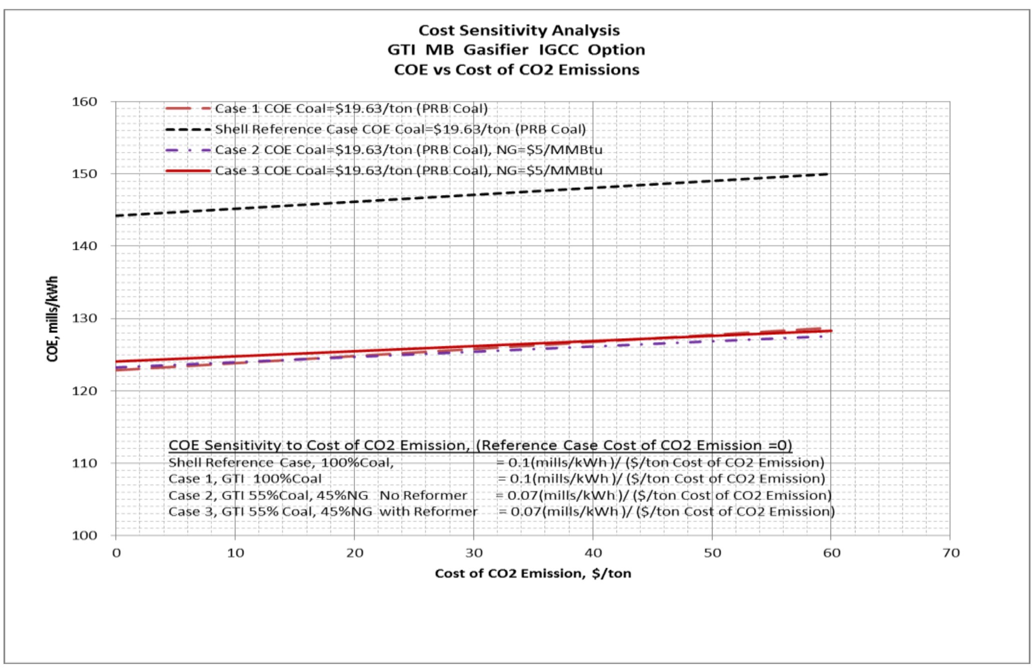


Figure 14. COE as a Function of CO2 Price for Shell and all HMB IGCC Cases

Techno-Economic Analysis – Fischer-Tropsch Configuration

Table 13 shows the TEA results for the HMB Fischer-Tropsch cases. Cases considered assumed a 55/45 coal/natural gas blend. The first case assumed direct raw gas reforming. The second gas considered parallel indirect reforming, and the third case assumed series indirect reforming. In all HMB FT cases, the cost of power (COP) based on diesel production was equal to or 1 to 2% higher than the Shell baseline case. The HMB cases all had lower fixed and variable operating costs compared with the Shell case, but they all had significantly higher fuel costs because gas has a higher price than powder river basin coal.

Table 13. Fischer-Tropsch TEA Results for Shell and HMB Gasification Plants

Case	Shell Gasifier Benchmark	Case 1FT Direct Reforming	Case 2FT Parallel Indirect Reforming	Case 3FT Seriesl Indirect Reforming
Capacity Factor (CF), %	90	90	90	90
Net Power Generation, MWe	17	29	11	1
2011 Capital Cost, $MM				
Total Plant Cost, $MM	7,327	6,156	7,350	6,924
Total Overniaht Cost. $MM	9.014	7.599	9.123	8.692
2011 Operating Cost, $MM/yr				
Fixed Operating Costs	*265*	*228*	*266*	*252*
Variable Operating Costs @ 90% CF	*192*	*149*	*175*	*169*
Fuel Costs @ 90% CF, Coal @$68.6/ton	*518.3*	*349.9*	*466.5*	*291.6*
NG @ $5.17/MMBtu	*0.0*	*505.0*	*191.0*	*412.0*
COP FT Diesel, excl CO2 TS&M, $/bbl FT diesel	202	202	212	207
COP FT Naphtha, excl CO2 TS&M, $/bbl FT Naphtha	141	141	148	144
COP FT ECO, excl CO2 TS&M, $/bbl ECO	157	157	165	160
COP FT EPD, excl CO2 TS&M, $/bbl EPD	196	196	206	200

The total plant cost summary in Table 14 shows the HMB cases are all equal to or lower than the Shell baseline case. The conclusions of this analysis are that 1) selecting the proper plant layout and method of raw syngas reforming are crucial, and 2) benefits of the HMB gasification process are dependent on the cost of coal and natural gas.

Laboratory HMB Testing

Testing results for the oxy-natural gas-coal tests are summarized in Figures 15-17 and Table 15. All tests were conducted in the HMB gasification test unit described above. Results are summarized for clarity in a series of graphs. The graphs show data from the different HMB tests as independent operating variables were changed. Independent variables included fuel gas and coal rates, steam or no steam, and oxygen rate (to adjust oxygen to fuel ratio). Several tests were also conducted with different coal injection locations above the molten slag bed.

Table 14. Fischer-Tropsch Total Plant Summary for Shell and HMB Gasification Plants

Code of Accounts	TOTAL PLANT COST, 2011 $MM	Shell Gasifier Benchmark	Case 1FT Direct Reforming	Case 2FT Parallel Indirect Reforming	Case 3FT Seriesl Indirect Reforming
1	COAL & SORBENT HANDLING	102.7	96.2	96.2	71.9
2	COAL & SORBENT PREP & FEED	548.7	511.9	511.9	375.4
3	FEEDWATER & MISC BOP SYSTEMS	92.2	67.5	67.5	61.8
4	GASIFIER & ACCESSORIES	3351.9	3619.2	3453.0	3538.8
5	GAS CLEANUP & PIPING	1172.2	1188.8	1188.8	1080.6
5AA	FT SYNTHESIS AND PRODUCT UPGRADE	975.2	962.8	962.8	932.9
5B.2	CO2 Compression & Drying	80.8	51.9	51.9	58.8
6	COMBUSTION TURBINE/ACCESSORIES	223.3	89.5	89.5	89.5
7	HRSG, DUCTING & STACK	74.1	79.1	79.1	72.8
8	STEAM TURBINE GENERATOR	241.5	257.7	257.7	237.0
9	COOLING WATER SYSTEM	75.2	60.3	60.3	66.3
10	ASH/SPENT SORBENT HANDLING SYS	112.1	97.5	97.5	72.7
11	ACCESSORY ELECTRIC PLANT	146.1	137.4	137.4	136.3
12	INSTRUMENTATION & CONTROL	37.0	36.4	36.4	36.4
13	IMPROVEMENTS TO SITE	51.8	52.0	51.9	51.8
14	BUILDINGS & STRUCTURES	42.3	41.4	41.3	41.3
	TOTAL TPC	7,327.4	7,349.6	7,183.2	6,924.4

The first series of HMB gasification tests was conducted with natural gas and oxygen sent to the burner and coal injected above the molten slag bed. In these tests, there was no steam or syngas injection. Data in Figure 15 shows that syngas quality, measured as higher H2/CO ratio and higher percentage CO in the product syngas, improved as the O2 to fuel ratio decreased and as the fraction of fuel as coal increased.

With a constant level of 49% coal on a fuel energy basis (the remainder coming from natural gas), the H2/CO ratio was found to be 0.52 to 0.60. Also with a constant 49% fuel value from coal, the percent CO in the product syngas increased with lower O2 to greater than 30% for an O2/natural gas ratio of 1.8.

Further tests with natural gas and no steam were made varying the coal fraction of the inlet fuel value. These tests found that increasing the fuel value in the form of coal from 49% to 56% increased the H2/CO ratio and the percent CO in product syngas. At an oxygen to natural gas ratio of 1.8, an increase in coal fuel value from 49% to 56%, the higher coal fuel value increase H2/CO ratio from 0.55 to 0.60 and increase the percent CO from 30% to 38%. Lower O2/natural gas ratio of 1.7 with 56% energy from coal continued the observed trend by increasing the H2/CO ratio to 0.62 and increasing the percent product syngas CO to 51%.

The second set of HMB gasification tests focused on adding steam to the coal-oxygen-natural gas process input streams. Results of these gasification conditions are summarized in Figure 16. The graphs show data from the different HMB test operating conditions. Adding steam increased the syngas quality. With 57% coal fuel input and an O2 to natural gas ratio of 1.7, steam addition of 0.25 standard cubic feet of steam per standard cubic foot of natural gas increased the H2/CO ratio from 0.5 to 0.7. The percent CO in the product syngas remained nearly constant, decreasing from 35% to 32%. The higher H2/CO ratio and constant CO concentration are confirmation of higher syngas quality.

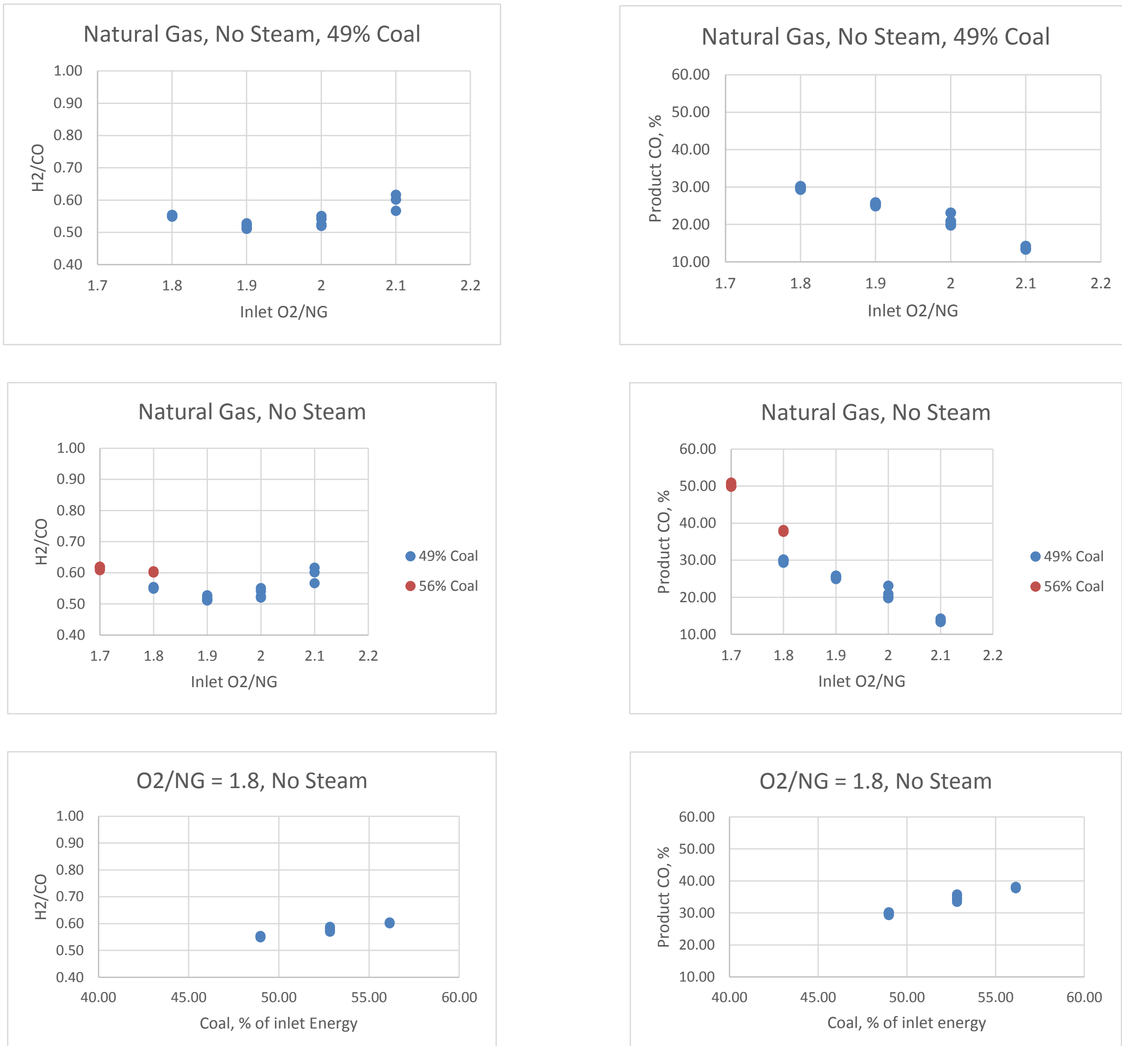


Figure 15. HMB Product Syngas From Tests Using Coal, Natural Gas, and No Steam

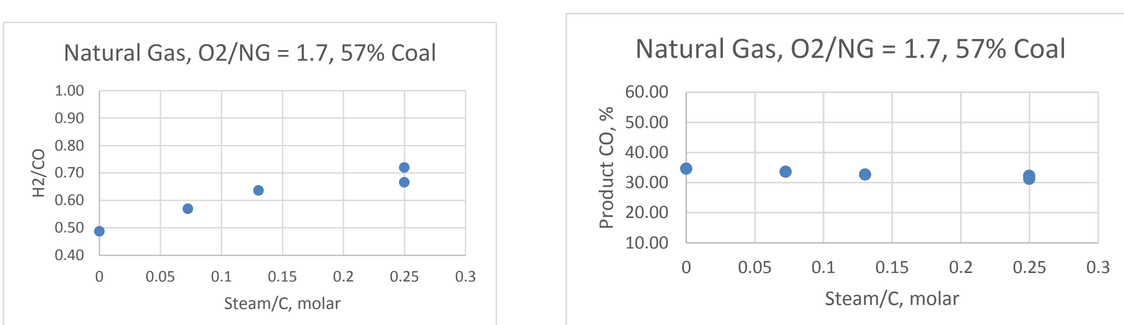


Figure 16. HMB Product Syngas From Tests Using Coal, Natural Gas, and Steam

The HMB gasification process recovers excess heat from the product syngas by combining feed natural gas with product syngas, partially reforming the combined gas stream, and sending the reformed syngas to the gasifier to be fired under reducing conditions in the oxy-gas burner. Two syngas compositions were selected to be representative of the HMB reformed syngas. The compositions of syngas 1 and syngas 2 are presented in Table 15.

Table 15. Syngas Compositions, dry basis

Component, vol%	Syngas 1	Syngas 2
CH4	24	24
H2	38	24
CO	38	52

Data in Figure 17 shows that switching feed natural gas to syngas increases the product H2/CO ratio and has a small impact on the percent CO in the product syngas. All syngas tests had an inlet steam/C ratio of 0.25. For an inlet combustion stoichiometry of 0.85, syngas with higher H2 content was found to produce product syngas with higher H2/CO ratio. Syngas 1 syngas generated a higher H2/CO ratio in the product gas than in the feed gas. Syngas 2, however, generated a lower H2/CO ratio in the product gas than in the feed gas.

The highest H2/CO ratio achieved in HMB gasification testing was 0.78. This was achieved with syngas 1 with coal introduced through a longer feed tube. The 12 inch coal feed tube extension allowed coal to be introduced closer to the surface of the molten slag bed. Data shows that the longer feed tube led to higher quality syngas with high H2/CO ratios.

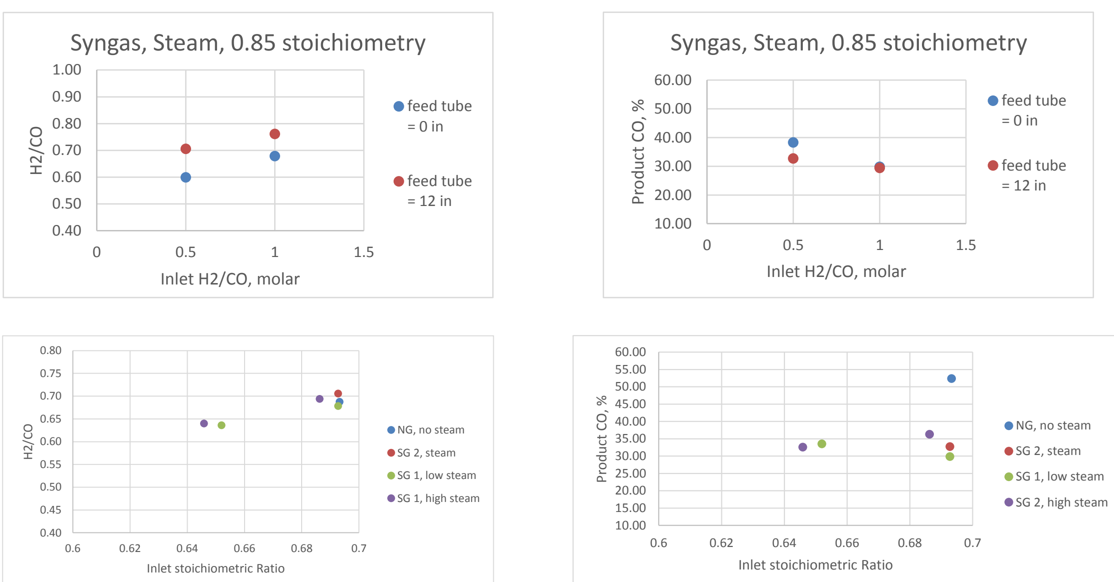


Figure 17. HMB Product Syngas From Tests Using Coal, Syngas, and Steam

Conclusion

The hybrid molten bed gasification process techno-economic analyses found that the HMB process is both technically and economically attractive compared with the Shell entrained flow gasification process. In IGCC configuration, HMB gasification provides both efficiency and cost benefits. In Fischer-Tropsch configuration, HMB shows small benefits, primarily because even at current low natural gas prices, natural gas is more expensive than coal on an energy cost basis. HMB gasification was found in the TEA to improve the overall IGCC economics as compared to the coal only Shell gasification process.

The non-recuperative HMB coal feed only IGCC operation case 1 found no efficiency difference vs Shell IGCC - 31.2% vs 31.2%. However, the cost of the HMB gasifier is lower vs Shell - $400 MM vs $750 MM. Overall, there is a lower COE - 122.8 mills/kWh vs 144.8 mills/kWh for the Shell baseline case.

For HMB IGCC recuperative operation with coal/natural gas co-feed and no reformer (Case 2) had an efficiency improvement of 2.2% (33.4% vs. 31.2%) from steam and natural gas preheats. This case had a higher COE – 125.0 mills/kWh vs 122.8 mills/kWh for the Shell case due to higher fuel cost.

For HMB IGCC recuperative operation with coal/natural gas co-feed with an external reformer (Case 3) additional efficiency improvement of 0.8% vs Case 2 (35.2% vs. 33.4%) is realized by heat recuperation through an external steam reformer. The COE is slightly higher - 125.8 mills/kWh vs 125.0 mills/kWh for Case 2 due to the added cost of a steam reformer.

The techno-economic analysis found the HMB gasification scheme for a FT CNTL plant is lower in capital cost compared to the Shell gasification based FT CNTL plant. COP for HMB direct reforming with coal/natural gas co-feed is $202/bbl of FT diesel. COP for a Shell gasification FT CNTL plant is the same at $202/bbl of FT diesel. This is explained by the higher fuel cost with the NMB cases having a 55.45 coal/gas blend and gas cost exceeding coal cost. The TEA work found that HMB gasification with indirect reforming (using an external steam reformer) has a higher cost due to the high cost of the external steam reformer. The HMB COP's are $212 and $287/bbl of FT diesel for the parallel and the series indirect reforming options.

Operationally, the HMB process proved to be robust and easy to operate. The burner was stable over the full oxygen to fuel firing range (0.8 to 1.05 of fuel gas stoichiometry) and with all fuel gases (natural gas and two syngas compositions), with steam, and without steam. The lower Btu content of the syngases presented no combustion difficulties.

The molten bed was stable throughout testing. The molten bed was easily established as a bed of molten glass. As the composition changed from glass cullet to cullet with slag, no instabilities were encountered. The bed temperature and product syngas temperature remained stable throughout testing, demonstrating that the bed serves as a good heat sink for the gasification process. Product syngas temperature measured above the bed was stable at ~1600ºF.

All originally planned testing was completed. The team did not achieve the target of a syngas with H2/CO ratio of 1.5 to 2.0. The highest H2/CO ratios achieved were between 0.7 and 0.78. This is higher than expected with gasification alone, but well below the HMB target. There are three primary directions needed to increase the HMB H2/CO ratio.

- First, temperature must be higher. Product syngas should be in the range of 2500°F, much higher than the 1600°F found in testing. Thermodynamics strongly favors higher hydrogen production as temperature is increased.
- Second, more effort must be put into designing a system to inject coal directly into the molten bed. Introducing coal above the bed likely led to coal being carried into the product gas and not entering the bed. This was confirmed when a longer feeder tube was used and H2/CO ratio increased. Efforts must be placed into methods to make sure all coal enters into the molten slag bed which is at approximately 2500°F. The conditions used for testing were likely not sufficient to reach the target.
- Third, the stoichiometric ratio must be even lower. As the ratio of oxygen to fuel decreased, the H2/CO ratio of the project syngas increased. Still lower oxygen to fuel ratio should continue this trend and help lead to the desired higher H2/CO ratio.

Operations data at different temperatures and stoichiometric ratios along with data on different feeder tube length shows trends toward the desired H2/CO ratio when the three directions listed above are met. Future testing should focus on changing conditions to reach the desired syngas H2/CO ratio.

Further HMB process testing should be carried out over a wider range of conditions. Tests should be conducted with different coals, coal particle sizes, and biomass. Longer operating times should be used for more complete process evaluation. Direct solid fuel injection into the molten bed should be employed along with wider inlet O2/fuel and inlet steam/carbon ratios. Extended testing should also cover higher pressure and different bed temperatures.

The recovery of process heat through product syngas reforming of feed natural gas is a promising route to higher gasification efficiency. Other gasification approaches may also benefit from the partial reforming of fuel gas. Further study of gasifier product gas reforming applied to HMB and other types of gasifiers may offer a way to improve the efficiency of several gasification technologies.

REFERENCES

None

AOI3 Hybrid Molten-Bed Gasifier for Production of High Hydrogen Syngas

Techno-Economic Analysis
IGCC Power Production Option

Submitted to

GTI / DOE NETL

By

Nexant

Contract No: DE-FE0009686

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1. INTRODUCTION

1.1. Background

Under the Department of Energy (DOE) Funding Opportunity Announcement (FOA) Number: DE-FOA-0000784, entitled "*Advanced Gasification Technologies Development and Gasification Scoping Studies for Innovative Initiatives*", Gas Technology Institute (GTI) is developing an innovative hybrid molten bed (HMB) gasification process to produce high-hydrogen syngas with hydrogen to carbon monoxide ratio (H_2/CO) from one to greater than six.

1.2. Study Objectives

This study will analyze an IGCC power plant with CO_2 capture that utilizes GTI's hybrid molten bed (HMB) gasification process. A technology and economic analysis study is required as a deliverable in the project Statement of Project Objectives.

2. IGCC DESIGN BASIS

2.1. Design References

NETL's *"Cost and Performance Baseline for Fossil Energy Plants Studies*" referred to as *"*Baseline Studies"[1] contained a comprehensive set of design basis and economic evaluation assumptions and criteria. These will be served as a reference for the purpose of the current study. DE-FOA-0000784 ATTACHMENT 2 also listed the following Baseline Studies references:

1. "*Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity* (Original Issue Date, May 2007), NETL Report No. 2010/1397, Revision 2, August 2010" -- (NETL Report 1397)
2. "*Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity: IGCC Cases*, NETL Report No. 2010/1399, May 2011" -- (NETL Report 1399)

The following recommended QGESS reports are also used to provide consistent design basis for feedstock and equipment specifications, and cost estimation methodology:

3. "*Detailed Coal Specifications*, NETL Report No. 401/012111, January 2012" -- (NETL Report 401/012111)
4. "*Process Modeling Design Parameters*, NETL Report No. 341/081911, January 2012" -- (NETL Report 341/081911)
5. "*Specification for Selected Feedstocks*, NETL Report No. 341/011812, January 2012" --- (NETL Report 341/011812)
6. "*CO_2 Impurity Design Parameters*, NETL Report No. 341/011212, August 2013" --- (NETL Report 341/011212)

NETL Report 1399 provides reference costs and economic evaluation guidelines. Additionally, the following reports also serve as reference sources for the economic evaluation reference in this study.

7. "*Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases*, August 2012, DOE/NETL-341/082312"-- (NETL Report 341/082312)
8. NETL's Series of Quality Guidelines for Energy Systems Studies (QGESS):
    - "*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"
    - "*Fuel Prices for Selected Feedstocks in NETL Studies*, November 2012, DOE/NETL 341/11212"

[1] http://www.netl.doe.gov/energy-analyses/baseline_studies.html

2.2. GTI Hybrid Molten Bed (HMB) Gasifier

The GTI HMB gasification is a dual coal-natural gas fueled process currently under development by GTI. While the HMB gasifier concept is new, the technology is based on commercially proven process called submerged combustion melting (SCM) that is used to produce a number of industrial products in the same temperature range and with the same oxy-gas burners. The following is a conceptual description of the HMB gasifier. GTI will provide additional details into the technical development and operational aspects of the HMB gasifier in another report.

In this innovative gasifier, natural gas and oxygen are fired under partial oxidation conditions upward into a bed of molten coal slag. The heat and gases generated drive the gasification process. Evaporative cooling walls generate steam for the gasifier to raise H_2 to CO ratio and to increase process efficiency. The syngas H_2/CO ratio can be optimized for producing electricity by IGCC or liquid fuels by Fischer-Tropsch synthesis by varying coal and natural gas feed rates, and steam to natural gas ratio. Control of the coal, natural gas, oxygen, and steam and their ratios to the gasifier generates a syngas whose H_2/CO ratio can be tailored accordingly. For diesel or other liquid fuels production, it is possible to generate a syngas that requires no water-gas shift reactor downstream of the gasifier whereas for IGCC purposes, a hydrogen-rich syngas can be produced directly from this gasifier such that it greatly reduces the downstream water gas shift reaction.

The HMB gasifier contains a bed of molten slag at 2500-2700°F maintained by fuel-rich combustion of natural gas and oxygen fired directly into the molten bath. Coal charged to the gasification chamber from above gasifies while circulating in the molten slag utilizing the excess heat from the partial oxidation of the fuel gas. The gasifier walls are built of tube banks with a thin layer of castable refractory on the inside. A thin layer of frozen slag forms on the walls, protecting them from abrasion, a process demonstrated with many mineral melts in GTI submerged combustion melters. Water that passes through the wall of tube banks forms steam. That steam is injected into the gasifier, providing additional reactant for steam carbon reactions while recuperating heat lost from the gasifier walls back to the gasifier. This innovative method of recovering heat from hot syngas improves overall plant efficiency since there is no other need for this heat. This feature is possible only in a hybrid gasifier because coal-only gasifiers are not designed to accept fuel gas. GTI has a long working history in the areas of natural gas reforming, recuperative reforming, and solid-gaseous co-firing technologies and it has drawn upon this extensive experience in developing the HMB gasifier.

2.3. Case Configurations

2.3.1. Reference Case and GTI HMB Gasifier IGCC Plant Techno-Economic Analysis Cases

Case S1B from the DOE/NETL 1399 Baseline Studies (reference 2, section 2.1) was selected as the reference case for this analysis. It is a Shell SCGP gasifier based IGCC power plant with CO_2 capture.

Three GTI HMB Gasifier cases with variations in feed mix, gasifier configuration and heat integration schemes are selected for the IGCC Plant Techno-Economic Analysis. Table 2-1 summarized the configurations of each case. Schematic depictions of these cases are included in the simplified block flow diagrams in figures 2-1 to 2-4.

1. Reference Case – Shell SCGP Gasifier, 100% PRB Coal Feed
2. Case 1- GTI HMB Gasifier, 100% PRB Coal Feed
3. Case 2- GTI HMB Gasifier, 55% PRB Coal / 45% NG Feed
4. Case 3- GTI HMB Gasifier, 55% PRB Coal / 45% NG Feed with Steam Reformer

Table 2-1
Summary of IGCC Power Plant Cases

	Reference Case[1]	Case 1	Case 2	Case 3
Gasification Technology				
Shell Dry Feed Gasification (SCGP)	X			
GTI Molten Bed Gasifier (MBG)		X	X	Hybrid MBG (with Pre-Reformer)
Feed Mix	100% Coal	100% Coal	55% Coal / 45% NG	55% Coal / 45% NG
Coal Type	PRB	PRB	PRB	PRB
Steam to Gasifier			X	
Syngas Cooling				
Recycle gas quench	X		X	
Water quench	X	X		
Steam Generation	X	X	X	X
Feed Preheat			X	X
Reformer Reaction Heat				X
Syngas Cleanup				
AGR - Selexol[2]	X	X	X	X
Water Gas Shift				
High & Low Temperature WGS	X	X	X	X
Gas Turbine				
Advanced Turbines	X	X	X	X
CO_2 Purification and Compression	X	X	X	X

1 Nexant's Simualtion of the DOE S1B reference case.

2 Additional trace contaminant cleanup such as mercury removal will be included as defined by DOE/NETL baseline studies.

Figure 2-1
Reference Case: Simplified BFD - Shell SCGP Gasifier IGCC Plant with 100% PRB Coal Feed

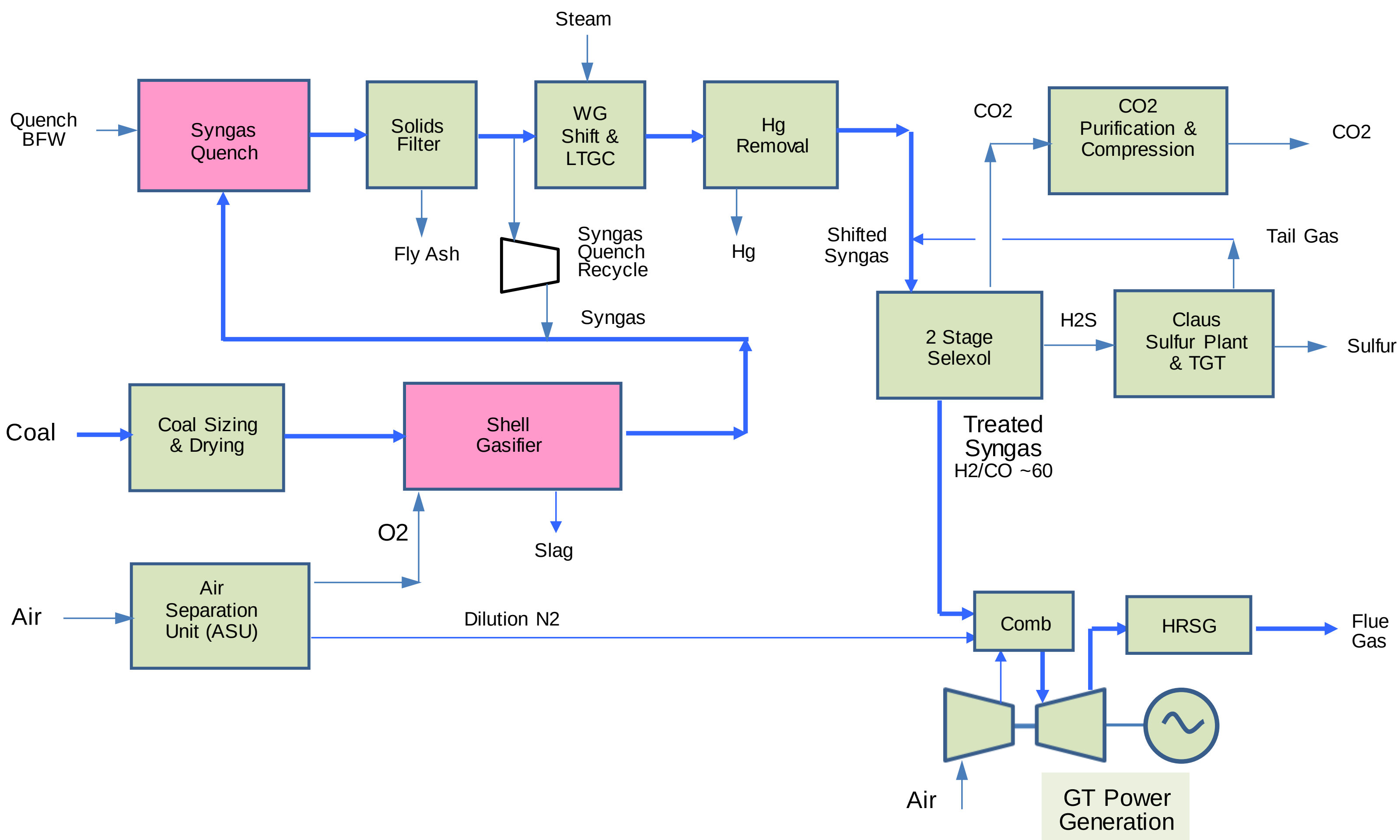


Figure 2-2
Case 1: Simplified BFD - GTI HMB Gasifier IGCC Plant with 100% Coal Feed and CO_2 Capture

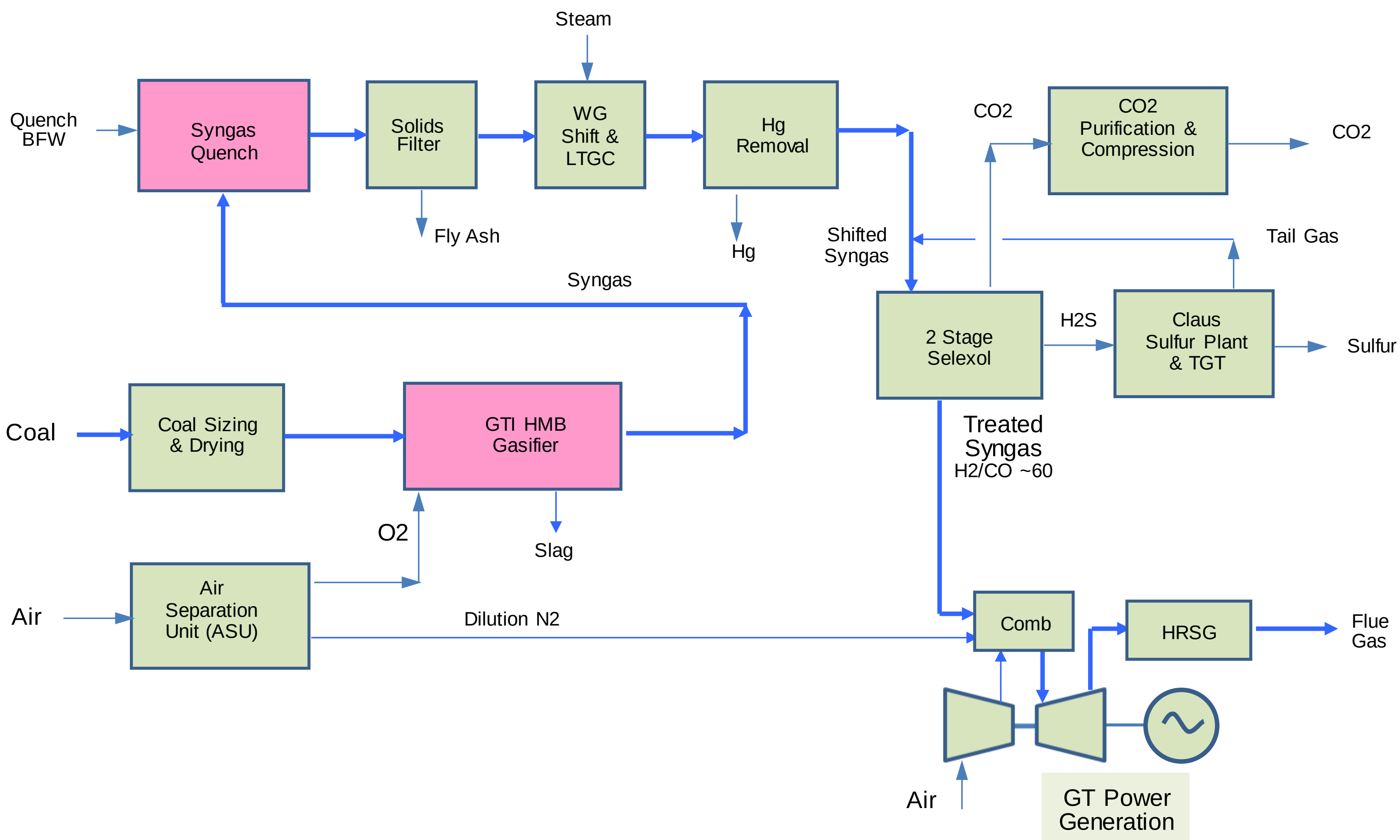


Figure 2-3
Case 2: Simplified BFD - GTI HMB Gasifier IGCC Plant with 55% Coal / 45% NG Feed and CO_2 Capture

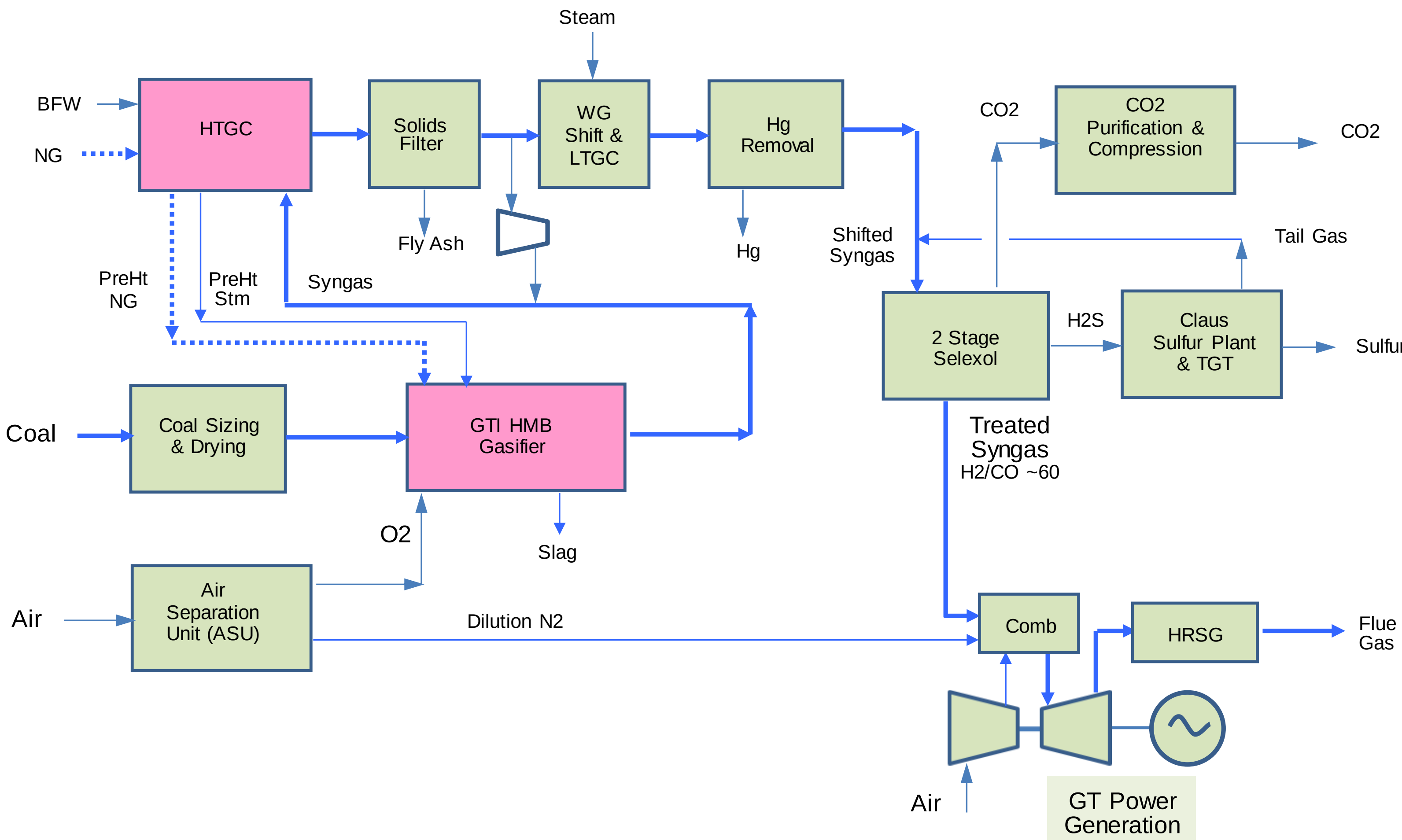
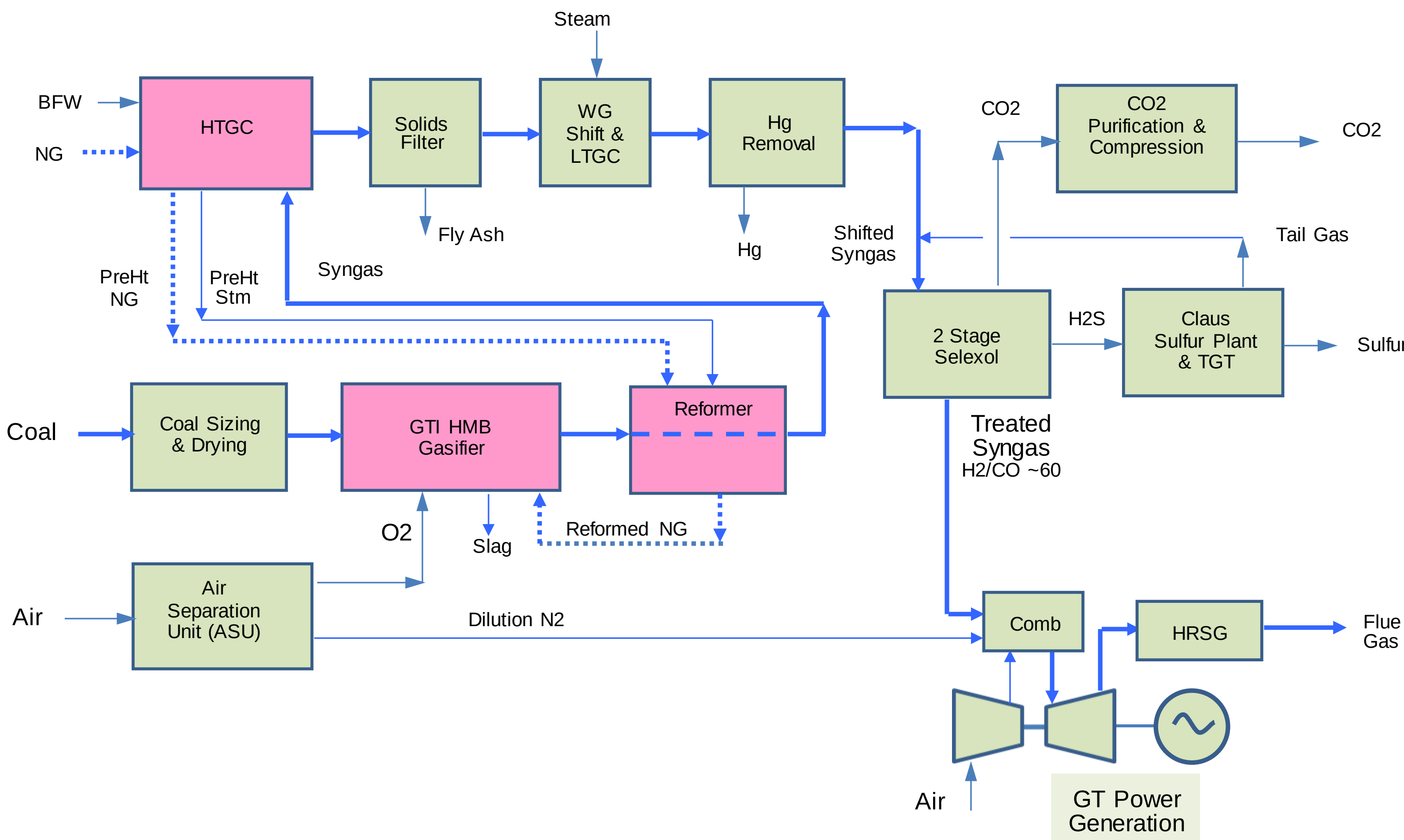


Figure 2-4
Case 3: Simplified BFD – Hybrid GTI HMB Gasifier IGCC Plant with 55% Coal / 45% NG Feed, Reformer and CO_2 Capture



2.3.2. Reference IGCC Power Plant with CO_2 Capture

The Reference IGCC power plant selected for the GTI Molten Bed (MB) techno-economic analysis is Case S1B from the NETL 1399 Baseline Study (reference 2 in section 2.1). A simplified Block Flow Diagram (BFD) for the Reference plant is shown in Figure 2-1.

Case S1B utilizes Shell gasification technology (SCGP) for syngas production and advanced GE 7F-turbines for power generation. Shell's gasification technology has been proven on a commercial scale and is considered technologically matured. Hence, its overall performance and cost can be estimated at a high confidence level.

The reference IGCC power plant is a 100% 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 MW each for a total of 430 MW at the Montana site's elevation. The power plant is equipped with a heat recovery steam generator (HRSG) and steam turbines to maximize power recovery. It is designed to capture CO_2 equivalent to 90% of the raw syngas' carbon content using the double-stage Selexol process. The nominal net IGCC power export capacity after accounting for the auxiliary loads which include CO_2 capture and compression is 460 MWe.

In order to achieve the 90% CO_2 removal target and maintain the same syngas heat content (Btu/SCF) to the GT, the raw syngas must be converted to hydrogen-rich syngas by the water gas shift (WGS) reaction. The shifted hydrogen-rich syngas has a H_2/CO ratio of ~60 compared to the raw syngas H_2/CO of 0.4. Steam for WGS is provided partly by vaporizing quench water during SG cooling and partly by saturating the water scrubber overhead gas. The balance of the WGS steam requirement is provided by steam addition to the WGS feed gas.

The WGS catalyst also hydrolyzes the COS to H_2S for capture in the AGR. The recovered H_2S is converted into elemental sulfur in the Claus plant.

The Reference IGCC power plant consists of the following major blocks:

- Coal Handling
- Coal Prep, Drying & Feed
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- Shell SCGP Gasifier System
- Syngas Cooling (Syngas Recycle Quench, Water Quench, Scrubbing, Steam Generation)
- Gas Cleaning (Filters, WGS, Hg Removal & AGR)
- CO_2 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
- Electrical Distribution

The IGCC plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 80 percent or 7,000 hours/year at full capacity.

2.3.3. Case 1: IGCC with CO_2 Capture - GTI Hybrid MB Gasifier with 100% Coal Feed

Case 1 IGCC power plant is similarly configured as the Reference Case except the Shell SCGP gasifier in the Reference Case is replaced by the GTI's MB gasifier. The GTI MB gasifier does not have the integral gas quench at the gasifier outlet which is part of the Shell SCGP gasifier design. A simplified block flow diagram for Case 1 IGCC plant is shown in Figure 2-2.

Like the Reference SCGP IGCC case, the Case 1 IGCC power plant is a 100% 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 MW each and includes a HRSG and steam turbines to recover waste heat from the GT flue gas for power generation. The double-stage Selexol process will capture 90% of the raw syngas' carbon content. The nominal net IGCC power export capacity after accounting for the auxiliary loads which include CO_2 capture and compression is 454 MWe.

The Case 1 IGCC power plant consists of the following major blocks. Differences between the Case 1 IGCC plant and the reference SCGP plant are in bold and italicized.

- Coal Handling
- Coal Prep, Drying & Feed
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- GTI HMB Gasifier System
- Syngas Cooling (Water Quench Only, Scrubbing, Steam Generation)
- Gas Cleaning (Filters, WGS, Hg Removal & AGR)
- CO_2 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
- Electrical Distribution

The IGCC plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 80 percent or 7,000 hours/year at full capacity.

2.3.4. Case 2: IGCC with CO_2 Capture - GTI Hybrid MB Gasifier with 55% Coal / 45% Natural Gas Mixed Feed

The Case 2 IGCC power plant is configured to use the GTI Hybrid Molten Bed (HMB) Gasifier designed for 55% Montana PRB coal / 45% natural gas co-feed. It is also designed to generate enough hydrogen-rich fuel gas to fill the two advanced 215 MW GE 7F-turbines and includes a HRSG and steam turbines to recover waste heat from the GT flue gas to maximize power generation. It is designed to capture CO_2 equivalent to 90% of the raw syngas' carbon content using the double-stage Selexol process. The nominal net IGCC power export capacity after accounting for the auxiliary loads which include CO_2 capture and compression is 510 MWe. A simplified block flow diagram for Case 2 IGCC plant is shown in Figure 2-3.

Case 2 syngas cooling/heat integration is optimized for high temperature natural gas and steam feed preheat and high pressure steam generation and superheat. With the high temperature feed preheat requirements it was deemed advantageous to maximize high level syngas heat recovery instead of using water quench for syngas cooling with the goal of improving the overall plant efficiency. In this heat integration scheme, hot syngas exiting the HMB gasifier at 2600°F is first quenched with cold recycled syngas to below the PRB coal ash fusion temperature of 2,238°F to prevent the deposition of molten ash in the downstream equipment. The quenched syngas at ~2,100°F provides the required duties and temperature driving force to achieve all of the following:

- Preheating the natural gas feed to 900°F
- Preheating the gasifier steam to 1200°F
- Generating superheated (1000°F) high pressure steam for the steam turbines.

In order to achieve the 90% CO_2 removal target, the raw syngas from the GTI MB gasifier (H_2/CO ratio of 1.3) is converted to hydrogen-rich syngas by Water Gas Shift (WGS) reactors. The shifted hydrogen-rich syngas has a H_2/CO ratio of ~66. Steam for WGS is provided partly by vaporizing quench water during SG cooling and partly by saturating the water scrubber overhead gas. The balance of the WGS steam requirement is provided by steam addition to the WGS feed gas.

The Case 2 IGCC power plant consists of the following major blocks. Differences between the Case 2 IGCC plant and the reference SCGP plant are in bold and italicized.:

- Coal Handling
- Coal Prep, Drying & Feed
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- GTI HMB Gasifier System
- Syngas Cooling (Syngas Recycle Quench, Water Quench, Scrubbing, Steam Generation)
- Gas Cleaning (Filters, WGS, Hg Removal & AGR)
- CO_2 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
- Electrical Distribution

2.3.5. Case 3: IGCC with CO_2 Capture - GTI Hybrid HMB Gasifier with Steam Reformer and with 55% Coal/45% Natural Gas Mixed Feed

The Case 3 IGCC power plant is configured to produce syngas for power generation by gasifying Montana PRB coal in the GTI Molten Bed gasifier and by reforming natural gas in an external steam reformer. The design feed mix is 55% Montana PRB coal and 45% natural gas on a HHV basis.

Natural gas is reformed with steam at 1,500°F in the external steam reformer. The reforming duty is provided to the reformer by heat exchange with the 2,600°F gasifier syngas. The reformer syngas contains ~10 mol% of unreacted CH_4 and has a H_2/CO mole ratio of 6.4. The unreacted CH_4 is further converted to H_2 and CO in the gasifier to minimize carbon slippage and improve cold gas efficiency.

PRB coal is gasified with 99.5% oxygen from the ASU in the GTI Molten Bed gasifier to produce syngas. At the gasifier outlet temperature of 2,600°F, most of the unreacted CH_4 from the reformer syngas are converted to H_2 and CO. The concentration of CH_4 at the gasifier outlet is negligible (<100 ppm). The H_2/CO mol ratio of the gasifier syngas is 1.4.

As per the previous cases, the Case 3 IGCC power plant is designed to generate fuel gas to fill two advanced GE 7F-turbines. It also includes a HRSG and steam turbines to recover waste heat from the GT flue gas to maximize power generation. The double-stage Selexol process captures 90% of the CO_2. The nominal net IGCC power export capacity after accounting for the auxiliary loads which include CO_2 capture and compression is 483 MWe. Figure 2-4 shows a simplified block flow diagram for the Case 3 IGCC plant.

Case 3 syngas cooling/heat integration is optimized to primarily provide the steam reforming duty and the natural gas and steam feed preheat duties. The balance of the syngas cooling duty is available for high pressure steam generation and superheat.

The hot syngas exits the GTI MB gasifier at 2,600°F and heat exchanges with the steam reformer to provide the required reforming duty. It leaves the reformer at ~1,800°F. At this temperature, there is enough driving force to preheat the natural gas feed and reformer steam to 900°F and 1,200°F respectively. Superheated (1000°F) high pressure steam is also generated from syngas cooling.

The Case 3 IGCC power plant consists of the following major blocks. Differences between the Case 3 IGCC plant and the reference SCGP plant are in bold and italicized.:

- Coal Handling
- Coal Prep, Drying & Feed
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- GTI MB Gasifier /Steam Reformer System
- Syngas Cooling (Water Quench Only, Scrubbing, Steam Generation)
- Gas Cleaning (Filters, WGS, Hg Removal & AGR)
- CO_2 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
- Electrical Distribution

The IGCC plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 80 percent or 7,000 hours/year at full capacity.

2.4. Process Design Parameters

2.4.1. 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 and fed through to the Shell or GTI MB gasifier. The gasifiers will gasify enough dried PRB coal to produce sufficient syngas to fully load two advanced GE 7F turbines (rated nominally at 215 MW 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 1399 Baseline Study, this study assumes that all sulfur in the coal is converted in the gasifier and leaves with the syngas.

2.4.2. Natural Gas Properties

Natural gas feed to the IGCC power plants is shown in the following table. The GTI HMB gasifiers will gasify enough dried PRB coal and natural gas to produce sufficient syngas to fully load two advanced GE 7F turbines (rated nominally at 215 MW each) at the Montana site's elevation.

Natural Gas Composition & Heating Values

Component	Volume Percentage	
Methane, CH_4	93.1	
Ethane, C_2H_6	3.2	
Propane, C_3H_8	0.7	
n-Butane, C_4H_{10}	0.4	
Carbon Dioxide, CO_2	1.0	
Nitrogen, N_2	1.6	
Total	100.0	
	LHV	HHV
Btu/SCF	932	1,032
Btu/lb	20,410	22,600

2.4.3. Gasification Block Process Design Parameters

The process design parameters for the gasification block based on the GTI Hybrid Molten Bed gasifier are summarized in table 2-3. The gasification block includes the gasifier system and the gas cooling/heat recovery and gas cleanup and CO_2 recovery facilities.

Table 2-3
Gasification Block Process Design Parameters

Case	DOE/NETL-2010/1399 Case S1B	GTI Hybrid Molten Bed
Gasifier Technology	Shell (SCGP)	GTI (HMB)
Coal Energy Content (%)	100%	>50%
Gasifier Pressure, (psia)	615	As Required
O_2:Coal Ratio, kg O_2/kg dry coal	0.773	As Required
Carbon Conversion, %	99.5	by GTI
Gasifier Heat Removal by Steam Generation, % Feed HHV	2%	2%
Gasifier Heat Loss, % of Feed HHV	1%	0%
Syngas HHV at Gasifier Outlet, (Btu/scf)	281	33,100 lbmol/h Total CO+H_2 @ HMB outlet
Nominal Steam Cycle, (psig/°F/°F)	1,800/1,000/1,000	1,800/1,000/1,000

Case	**DOE/NETL-2010/1399 Case S1B**	**GTI Hybrid Molten Bed**
Condenser Pressure, (in Hg)	1.4	As Required
Combustion Turbine	2x Advanced F Class (Nominal 232 MW output each, reduced by elevation considerations)	2x Advanced F Class(215MW output each @ 3,400 feet elevation)
Oxidant	95 vol% Oxygen	Same
Coal	Subbituminous	Same
H_2S Separation	Selexol (1st Stage)	Same
Sulfur Removal, %	99.7	As Required
CO_2 Separation	Selexol (2nd Stage)	Same
CO_2 Removal, %	90	90
Sulfur Recovery	Claus Plant with Tail Gas Treatment / Elemental Sulfur	Same
Particulate Control	Cyclone, Candle Filter, Scrubber, and AGR Absorber	Same
Mercury Control	Carbon Bed	Same
NO_x Control	MNQC (LNB) and N_2 Dilution	Same

2.4.4. GTI Block Design Criteria

GTI Block is designed as an integral part of the advanced IGCC plant with 90% feed carbon recovery as CO_2 for sequestration into saline reservoirs. It includes the following major gasification and syngas cleanup related systems:

- Feed Pressurization System
- HMB Gasifier
- Reformer
- Syngas Cooling and Reforming Steam Generation
- CO_2 Compression and Purification Facilities
- O_2 Booster Compressor
- NG Booster Compressor and Preheat
- Reforming Steam Preheat

Nexant envisions establishing Case S1B, from the ‘Baseline Studies Report’ as the reference IGCC case for our techno-economic study of the GTI HMB gasifier. Case S1B is a Shell gasifier based IGCC power plant with CO_2 capture.

2.4.5. Non-GTI Block Design and Criteria

The Non-GTI Block (NGB) includes the systems common to both the conventional and GTI HMB IGCC plants, which are not directly related to the advanced coal gasification and syngas cleanup systems. Apart from being of different capacities, these systems are expected to have nearly identical flow schemes as the corresponding conventional IGCC with CO_2 capture cases reference case S1B from NETL Report 1399. Due to the similarity in designs between these systems that are common to both the advanced and conventional IGCC cases, the Non-GTI Block systems costs will be scaled based on capacity factors given in the QGESS *Capital Cost Scaling Methodology* document for the advanced IGCC plant wherever possible.

Process modeling for the NGB systems will be carried out, to the maximum extent possible, in accordance with guidelines from the Baseline Study NETL 1399 and QGESS *Process Modeling Design Parameters* documents. This is used mainly to determine the utilities consumption or power generation rates of the NGB systems in order to evaluate the overall IGCC plant efficiency.

2.4.6. Gas Turbine Design Criteria

For this study, GTI HMB gasification system produces syngas for two advanced F-Class combustion turbines (CT). At ISO conditions, gross turbine power, as measured prior to the generator terminals, is 232 MWe each for a total of 464 MWe. Turbine output is reduced at 3,400 feet elevation for the study site because the compressor capacity on a mass flow basis is reduced due to the reduced ambient air density. Nexant's design will use the gas turbine design condition of 215 MWe each for a total of 430 MWe according to NETL baseline study for capture cases at Montana site.

The power plant is also equipped with a heat recovery steam generator (HRSG) and steam turbine to generate additional power from waste heat from the flue gas. Adding in the steam turbine power and net of auxiliary loads, the plant's net capacity will have a nominal range of between 450 and 500 MWe.

Hot combustion products are expanded in the three-stage turbine-expander. Given the assumed ambient conditions, back-end loss, and HRSG pressure drop, the CT exhaust temperature is nominally 1,050°F for capture cases.

2.4.7. Steam Cycle Design Criteria

For this study, a GateCycle™ model of the steam cycle is developed and calibrated against the Baseline Report 1399 Case S1B IGCC power plant steam cycle characteristics. The GateCycle™ model will be re-run to estimate the power plant STG performance for the different optimizations expected for the current project. Selected steam cycle flows and operating conditions for developing and bench-marking the GateCycle™ model are listed below:

Steam Conditions for IGCC Technologies

Main Steam Pressure, psig	1,800
Main Steam Temperature, °F	1,000 (Range 950-1075)
Reheat Steam Temperature, °F	1,000 (Range 950-1075)

2.4.8. Cooling Water

It is assumed that GTI HMB/IGCC power plant utilizes a mechanical draft, evaporative recirculating wet cooling tower, and all process blowdown streams are assumed to be treated and recycled to the cooling tower. According to the NETL "Process Modeling Design Parameters, Rev. January 17, 2012" QGESS reference, typical cooling tower approach temperatures are in the range of 8 – 20°F for the power plant applications. For the Montana location with ambient wet bulb temperature of 37°F, NETL systems studies use an approach to wet bulb of 11°F. Cooling water range is assumed to be 20°F. Cooling water from the cooling towers is thus available at the following conditions:

- Maximum supply temperature, °F 48
- Maximum return temperature, °F 68
- Assumed maximum supply pressure, psia 60
- Assumed maximum cooler pressure drop, psi 10

Cooling tower makeup rate calculation is also specified by the same NETL QGESS, and is determined as followed:

- Evaporative losses = 0.8 percent of the circulating water flow rate per 10°F of range
- Drift losses = 0.001 percent of the circulating water flow rate
- Blowdown losses = Evaporative Losses / (Cycles of Concentration - 1)

  where cycles of concentration are a measure of water quality, and a mid-range value of 4 is chosen for this study

2.4.9. Air Separation Unit (ASU) Design Criteria

The air separation plant is designed to produce 95 mole percent O_2 for use in the gasifier. The plant is designed with two production trains, one for each gasifier. The air compressor is powered by an electric motor. Nitrogen is also recovered, compressed, and used for fuel gas dilution in the GT combustor.

Conventional cryogenic ASU will be used to produce the 95 mole percent purity oxygen for use in the GTI HMB gasification. The ASU will be designed for ambient air quality as shown in Table 2-4. Product oxygen composition is listed in Table 2-5 below. An oxygen compressor will be provided to boost the product oxygen pressure to that required to feed the GTI HMB gasifier. ASU performance and utility consumption will be pro-rated from the NETL Report 1399 design based on total oxygen production.

Table 2-4
Ambient Air Quality

Air composition based on published psychrometric data, mass %	
Argon	1.283
CO_2	0.050
O_2	23.049
N_2	75.220
Moisture	0.398
Total	100.00
Air Composition, mol%	
Argon	0.93
CO_2	0.03
O_2	20.81
N_2	77.59
Moisture	0.64
Total	100.00
Site Conditions:	
Ambient Pressure, psia	13
Design Ambient Temperature, Dry Bulb, °F	42
Design Ambient Temperature, Wet Bulb, °F	37
Design Ambient Relative Humidity, %	62

Table 2-5
Product Oxygen Quality

Analysis by Weight:	Volume %
N_2 O_2 Argon Total Vol%	1.78 95.04 3.18 100.00
Conditions before Booster Compression:	
Pressure, psia	125
Temperature, °F	90

2.4.10. Balance of Plant

Table 2-6
Balance of Plant

Fuel and Other Storage	
Coal	30 days
Slag	30 days
Sulfur	30 days
Sorbent	30 days
Plant Distribution Voltage	
Motors below 1 hp	110/220 volt
Motors between 1 hp and 250 hp	480 volt
Motors between 250 hp and 5,000 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam and CT Generators	24,000 volt
Grid Interconnection Voltage	345 kV

2.4.11. CO_2 Product Treating and Purification Design Criteria

For this study, recovered CO_2 is delivered at the battery limit (B/L), with specifications for saline reservoir sequestration listed in Table 2-7, per the NETL "CO_2 Impurities Design Parameters, Draft Report, August 23, 2013"QGESS reference.

Table 2-7
B/L CO_2 Pipeline Specifications[2,3]

B/L Pipeline Pressure, psia	2,215
B/L Pipeline Temperature, °F	95
Compositions:	
CO_2, vol% (Min)	95
N_2 + Ar, vol% (Max)	4
O_2, vol% (Max)	4
CH_4 + H_2, vol% (Max)	4
CO, ppmv (Max)	35
SO_2, ppmv (Max)	100
NO_x, ppmv (Max)	100
H_2O, ppmv (Max)	300

CO_2 compression facilities will be provided to boost the CO_2 product pressure to the required B/L requirement.

2.4.12. Water Supply and Waste Water

<u>Makeup Water</u>
The water supply is 50 percent from a local publicly owned treatment works (POTW) and 50 percent from groundwater, and is assumed to be in sufficient quantities to meet plant makeup

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[2] http://www.netl.doe.gov/energy-analyses/pubs/LR_IGCC_FR_20110511.pdf
[3] http://www.netl.doe.gov/energy-analyses/refshelf/PubDetails.aspx?Action=View&PubId=420

requirements. Makeup for potable, process, and de-ionized (DI) water is drawn from municipal sources.

Process Wastewater

Water associated with gasification activity and storm water that contacts equipment surfaces is collected and treated for discharge through a permitted discharge.

Sanitary Waste Disposal

Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site. Packaged plant was sized for 5.68 cubic meters per day (1,500 gallons per day)

Water Discharge

Most of the process wastewater is recycled to the cooling tower basin. Blowdown is treated for chloride and metals, and discharged.

2.4.13. Environmental/Emissions Requirements

The IGCC environment targets were established in the Electric Power Research Institute's (EPRI) design basis for their CoalFleet for Tomorrow Initiative, documented in the CoalFleet User Design Basis Specification for Coal-Based Integrated Gasification Combined Cycle (IGCC) Power Plants, EPRI, Palo Alto, CA, 2009. The design targets were established specifically for bituminous coal but apply to subbituminous case as well. The emissions requirements and limits for the reference IGCC power plant, as specified in NETL Report 1399, are listed below:

Table 2-8
IGCC Environmental Targets

Pollutant	Environmental Target	NSPS Limit
NOx	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 for the GTI HMB gasification IGCC process.

2.4.14. Overland Transportation Size Limitations

The site is listed to be landlocked with access by train and highway only. Maximum overland highway transportable dimension is assumed to be 100 feet long by 12 feet wide by 15 feet height (including carriage height). Maximum equipment height is 13.5 feet assuming using 1.5 feet height low boy carriage. Maximum overland highway transportable weight is 65 tons.

Maximum railway transportable dimension is assumed to be 100 feet long by 12 feet wide by 19 feet height (including railcar height). Maximum equipment height is 15 feet assuming using 4 feet height railcar. Maximum railway transportable weight is assumed to be 130 tons.

2.4.15. 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

2.5. 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 2,200 psig at IGCC battery limit and transported 50 miles for sequestration in a saline formation at a depth of 4,055 ft (Study scope limited to delivery at battery limit only)

2.6. 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), °F	42
▪ Wet bulb temperature (WBT), °F	37
▪ Ambient relative humidity, %	62

2.7. Capital Cost Estimation Methodology

2.7.1. General

For IGCC plants with CO_2 capture, the NETL 1399 Baseline Study provided a code of accounts grouped consisting of 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 Case 1, Case 2 and Case 3 GTI HMB gasifier based IGCC plants, except for the costs of the HMB gasifier and steam reformer, capital cost scaling following the guidelines and parameters described in the NETL *Capital Cost Scaling Methodology* document is used to perform the cost estimates for non-feed system related costs. 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 Baseline Study. Once these have been established, the capital cost can be estimated based on the revised capacity from the heat and material balances developed by Nexant.

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

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

Construction indirect cost are 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 1399 report.

2.7.2. Balance of Plant Capital Cost Estimate Criteria

For the rest of the systems that are not related to coal feeding, the capital cost estimates are developed based on the Case S1B Shell gasifier-based IGCC plant with CO_2 capture case in NETL 1399 Baseline Study.

For the these sections' subsystems, capital cost scaling following the guidelines and parameters described in the NETL *Capital Cost Scaling Methodology* document is used to perform the cost estimates, as described in Section 2.12.1.

Table 2-9 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 BOP subsystems, as laid out by the NETL *Capital Cost Scaling Methodology* document, are also shown in this table.

Table 2-9
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, Quench Column, Filters & Cyclones	Syngas Throughput
4.1a	Steam Reformer	Calculated
4.1b	Natural Gas Compression	Calculated
4.2	Syngas Heat Recovery	Calculated
4.3	ASU/Oxidant Compression	O_2 Production
4.4	Scrubber & Low Temperature Cooling	Syngas Flow
Acct No.	**Item/Description**	**Scaling Parameter**
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

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_2O Flow Rate
9.3	Circ Water System Auxiliaries	Circ H_2O Flow Rate
9.4	Circ Water Piping	Circ H_2O Flow Rate
9.5	Makeup Water System	Raw Water Makeup
9.6	Component Cooling Water System	Circ H_2O Flow Rate
9.9	Circ Water System Foundations	Circ H_2O 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
Acct No.	**Item/Description**	**Scaling Parameter**
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 Pump House	Circ H_2O 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

2.7.3. Home Office, Engineering Fees and Project/Process Contingencies

Engineering and Construction Management Fees and Home Office cost, project and process contingencies will be factored from the each subsystem's TFC. These are then added to the TFC to come up with the total project cost (TPC) of the system. Factors from Case S1B in the NETL 1399 Baseline Report will be used for this study.

2.7.4. Owner's Cost

Owner's cost is then added to TPC to come up with the total overnight cost (TOC) for the system. Owner's costs as defined in the NETL 1399 Baseline 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

- Initial Cost for Catalyst and Chemicals per design
    - Land Cost = $900,000 at 300 acres x $3,000/acre
    - Other Owner's Costs at 15% TPC
    - Financing Costs at 2.7% TPC

2.8. Operation & Maintenance Costs

The operation and maintenance (O&M) costs pertain to those charges associated with operating and maintaining the 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 are estimated based on 80% capacity factor.

2.8.1. Fixed Costs

Operating labor cost is 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, $/hour	$39.7
Length of work-week, hours	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

2.8.2. Variable Costs

The cost of consumables, including fuel, is 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 are evaluated similarly to the consumables.

The unit costs for major consumables and waste disposal will be selected from NETL 1399 Baseline Report, 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 Montana IGCC plant is $19.63/ton, per the QGESS *Fuel Prices for Selected Feedstocks in NETL Studies* document. The price of natural gas is $5.34/1000ft^3 ($5.17/MMBtu HHV) per QGESS.

2.8.3. CO_2 Transport and Storage Costs

As specified in DE-FOA-0000784 Attachment 2, CO_2 Transport and Storage (T&S) costs storage for the Montana IGCC plant location is $22/tonne. Per the TEA reporting requirements, the COEs will be reported both with and without the cost of CO_2 T&S.

2.9. Financial Modeling Basis

2.9.1. Cost of Electricity

The metrics used to evaluate overall financial performance are the cost of electricity (COE) for the IGCC plant. All costs are expressed in the "first-year-of-construction" year dollars, and the resulting COE is also expressed in "first-year-of-construction" year dollars.

The same financial modeling methodology is used for this study as per the NETL 1399 Baseline Study, which in turn is consistent with 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(OC_{FIX} and OC_{VAR}), capacity factor(CF) and net power generation (MWH), as shown in the equation below:

$$COE = \frac{\begin{matrix} first\ year \\ capital\ charge \end{matrix} + \begin{matrix} first\ year \\ fixed\ operating \\ costs \end{matrix} + \begin{matrix} first\ year \\ variable\ operating \\ costs \end{matrix}}{\begin{matrix} annual\ net\ megawatt\ hours \\ of\ power\ generated \end{matrix}}$$

$$COE = \frac{(CCF)(TOC) + OC_{FIX} + (CF)(OC_{VAR})}{(CF)(MWH)}$$

The capital charge factor (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 CO_2 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.

2.9.2. CO_2 Sales Price

As outlined in the TEA's reporting requirements, sensitivity analysis is to be done to determine the impact of CO_2 sales on IGCC COE. The varying parameter is the CO_2 sales price at the IGCC plant gate and is to range 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 CO_2 sales is shown below:

$$COE = Baseline\ COE\ - \frac{(CO_2\ Sales\ Price)\ \times annual\ tonnes\ of\ CO_2\ product}{annual\ net\ megawatt\ hours\ of\ power\ generated}$$

2.9.3. Cost of CO_2 Emissions

The TEA also requires sensitivity analysis on cost of CO_2 emissions to be performed. The varying parameter is the CO_2 emissions cost. The range of the emissions cost is between \$0/tonne (baseline case assuming no CO_2 emissions cost) and \$60/tonne.

The formula used to calculate the revised COE after taking into cost of CO_2 emissions is shown below:

$$COE = Baseline\ COE + \frac{(Cost\ of\ CO_2\ Emissions)\ \times annual\ tonnes\ of\ CO_2\ emitted}{annual\ net\ megawatt\ hours\ of\ power\ generated}$$

3. REFERENCE CASE: REFERENCE SHELL IGCC PLANT WITH CO_2 CAPTURE

3.1. Process Overview

The Reference IGCC power plant selected for the GTI Molten Bed (MB) techno-economic analysis is Case S1B from the NETL 1399 Baseline Study (reference 2 in section 2.1). A simplified Block Flow Diagram (BFD) for the Reference plant is shown in Figure 2-1.

Case S1B utilizes Shell gasification technology (SCGP) for syngas production and advanced GE 7F-turbines for power generation. Shell's gasification technology has been proven on a commercial scale and is considered technologically matured. Hence, its overall performance and cost can be estimated at a high confidence level.

The reference IGCC power plant is a 100% 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 MW each for a total of 430 MW at the Montana site's elevation. The power plant is equipped with a heat recovery steam generator (HRSG) and steam turbines to maximize power recovery. It is designed to capture CO_2 equivalent to 90% of the raw syngas' carbon content using the double-stage Selexol process. The nominal net IGCC power export capacity after accounting for the auxiliary loads which include CO_2 capture and compression is 460 MWe.

In order to achieve the 90% CO_2 removal target and maintain the same syngas heat content (Btu/SCF) to the GT, the raw syngas must be converted to hydrogen-rich syngas by the water gas shift (WGS) reaction. The shifted hydrogen-rich syngas has a H_2/CO ratio of ~60 compared to the raw syngas H_2/CO ratio of 0.4. Steam for WGS is provided partly by vaporizing quench water during SG cooling and partly by saturating the water scrubber overhead gas. The balance of the WGS steam requirement is provided by steam addition to the WGS feed gas.

The WGS catalyst also hydrolyzes the COS to H_2S for capture in the AGR. The recovered H_2S is converted into elemental sulfur in the Claus plant.

The Reference IGCC power plant is consisted of the following major blocks:

- Coal Handling
- Coal Prep, Drying & Feed
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- Shell SCGP Gasifier System
- Syngas Cooling (Quench, Scrubbing, Steam Generation)
- Gas Cleaning (Filters, WGS, Hg Removal & AGR)
- CO_2 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

- Electrical Distribution

The IGCC plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 80 percent or 7,000 hours/year at full capacity.

3.2. Process Description

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.

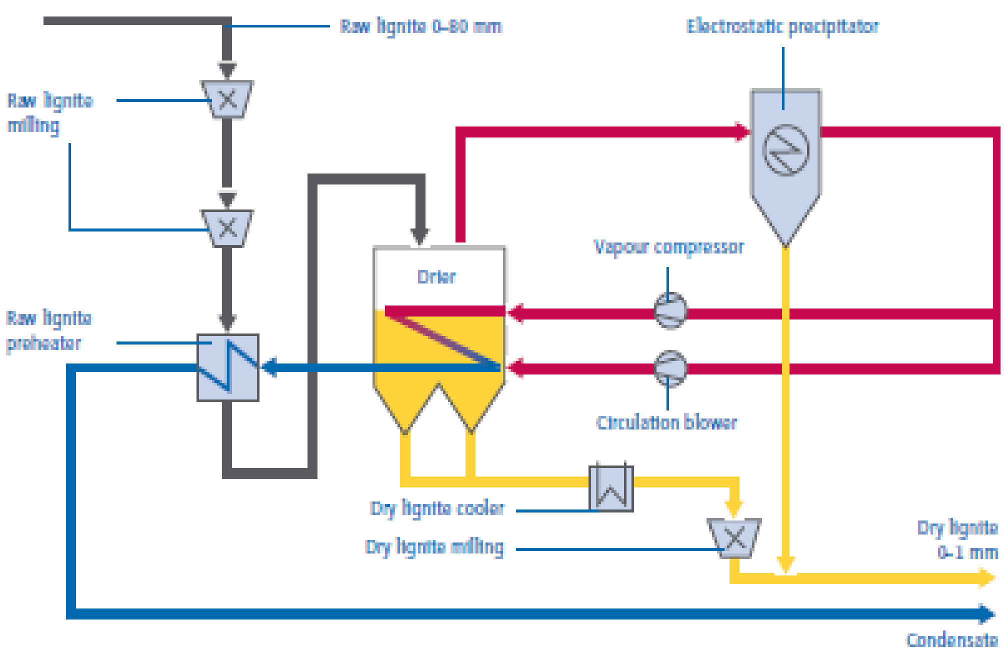
3.2.2. Coal Drying

Dry coal feed moisture content of 6% is used for the Shell entrained flow gasifier in Reference S1B case of the DOE/NETL 1399 report.

A paper presented by Shell in the Gasification Technology Conference was cited in the Reference S1B case as recommending drying subbituminous coal to 6% moisture before feeding 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 1399 report is the 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 below.

Figure 3-1
WTA Coal Drying Process Schematic



3.2.3. ASU

The Shell Reference case S1B utilizes an "elevated pressure" ASU in which the main air compressor discharge pressure is set at 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 NOx.

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	125	90
Diluent N_2	384	385
Transport N_2	815	387
ASU Vent	164	64

3.2.4. Gasification

Two trains of Shell dry feed gasifiers are used to process a total of 6,513 tons/day of as-received Montana PRB coal (5,552 tons/day of 6% dry coal). These gasifiers operate at 615 psia. Coal is gasified with 95% oxygen from the ASU to produce syngas containing H_2, CO, CO_2, H_2O, NO_x, SO_x and other products of coal gasification. The gasifier membrane wall is cooled by steam generation to create a protective ash layer over the membrane to maintain gasification temperature at ~2,600°F.

3.2.5. 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 lock hopper for dewatering. The water is clarified and reused as makeup to the water scrubber. The dried slag is stored for disposal.

3.2.6. Syngas Cooling & Particulate Filters

The raw syngas from the gasifier is quenched to below the ash melting point (~2,298°F) by cooled recycled syngas. It is then water quenched to 685°F before entering the ceramic particulate filters and cyclones. Any remaining particulate matters in the syngas will be removed by these filters and cyclones.

The filtered syngas is then cooled to 450°F through a series of steam generators generating steam for the steam cycle. The cooled syngas enters a water scrubber to remove any chlorides and remaining particulates. The scrubbed syngas is sent to WGS. Part of the scrubber bottoms is used for slag water bath makeup.

3.2.7. Water Gas Shift & COS Hydrolysis

In order to achieve the 90% CO_2 removal target and maintain the same syngas heat content (Btu/SCF) to the GT, the raw syngas must be converted to hydrogen-rich syngas by the water gas shift (WGS) reaction. The shifted syngas is hydrogen-rich and has a H_2/CO ratio of ~60 compared to the raw syngas H_2/CO ratio of 0.4.

The WGS reactors are located downstream of the water scrubber and ahead of the AGR. They contain sulfur tolerant shift catalysts for the WGS and COS hydrolysis reactions. There are two trains of WGS reactors with two reactor stages in series for the specified CO_2 capture requirements. Inter-stage cooling by steam generation is required to control the exothermic temperature rise in the reactors.

3.2.8. Mercury Removal

The shifted syngas from the WGS is cooled to 100°F and sent to a packed carbon bed vessel designed to treat the cooled syngas for mercury removal. One carbon bed vessel is required for each gasifier train. The beds are designed for superficial gas velocity of 1 ft/sec and design residence time of 20 seconds.

3.2.9. Selexol Acid Gas Removal (AGR)

H_2S and CO_2 are removed from the cooled syngas in a double stage Selexol AGR. H_2S is preferentially removed in the first stage absorber with CO_2 removal occurring in the second absorber stage. The treated syngas is reheated and sent to the gas turbine for power generation. To meet the environmental emission target of 0.0128 lb/MMBtu for SO_2, the H_2S rich stream from the first stage absorber is sent to the Claus sulfur plant where the H_2S is converted to elemental sulfur for recovery. The Claus plant tail gas is further treated in a tail gas treating unit (TGTU) to meet the SO_2 emission target. The CO_2 rich gas from the second stage absorber is sent to purification and compression.

3.2.10. Claus Plant

Since oxygen is available from the ASU, oxygen-blown Claus sulfur plant is selected for recovering sulfur from the process acid gas streams. The process streams include H_2S rich streams from the AGR, TGTU tail gas recycle and the SWS off-gas. The oxygen-blown Claus process can provide operating flexibility and lower cost than conventional Claus process.

The H_2S rich feed streams are first combusted in the Claus sulfur reaction furnace to form SO_2 which is then converted to elemental sulfur in the catalytic reactors. Three catalytic stages and a tail gas treating unit (TGTU) are required to achieve the sulfur recovery efficiency of 99.8%. A catalytic hydrogenation TGTU unit is used in this evaluation to be consistent with the Shell reference case.

3.2.11. CO_2 Compression and Dehydration

Raw CO_2 greater than 99% purity leaves the Selexol AGR plant the battery limit conditions of 150 psia and 60°F. It is compressed to supercritical condition of 2,215 psia using a multi-stage, intercooled compressor. The CO_2 stream is dehydrated to a dew point of -40°F at the appropriate inter-stage pressure using a thermal swing adsorptive dryer.

3.2.12. Gas Turbine

The gas turbine 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 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 is 215 MW at the Montana site location.

3.2.13. Steam Turbine and HRSG

The 1,050°F GT exhaust is cooled in the HRSG by generating HP, IP and LP steams for the steam turbines (ST) and process users. The cooled GT flue gas exits the HRSG at 270°F and is vented to the atmosphere through a stack. 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 send 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.

3.2.14. BOP

Raw Water System

Raw water system supplies cooling tower makeup, demineralizer water makeup, fire protection system water and potable water requirements. The water source is 50 percent from potable water and 50 percent from groundwater.

The demineralizer makeup system consists of two 100 percent trains, each with a 100% capacity activated carbon filter, primary cation and anion exchanger, mixed bed exchanger, recycle pump, and regeneration equipment. It provides demineralized water for HRSG BFW makeup.

The fire protection system provides pressurized water to the fire hydrants, hose stations and fire suppression sprinkler systems.

Accessory Electric Plant

The accessory electric plant is consisted 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.

Instrumentation and Control

A plant wide distributed control system (DCS) is provided.

3.3. Shell Reference case Performance

Nexant performed a benchmark of the reference Case S1B using in-house simulation programs The benchmarked performance is compared with the reference case in Table 3-2. The NETL 1399 Report provided a breakdown of the Case S1B (Shell Gasification-based IGCC with CO_2 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. Nexant provided its benchmarked version of the S1B IGCC plant's power generation and auxiliary loads. The power generation portion was calculated using Nexant's model of the S1B case, while the auxiliary loads were estimated by pro-rating from the relevant scaling parameters obtained from the heat and material balance developed by the same model.

Table 3-2 shows the power production and auxiliary load breakdown of the original DOE/NETL S1B case from the NETL 1399 Report (Case S1B) and the Nexant S1B Benchmark case 2a.

Table 3-2
Power Generation and Auxiliary Load Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	**Shell Reference Case (Bench Mark 2a)**	**Shell Reference Case (Case S1B)**
Gas Turbine Power	429,974	430,900
Steam Turbine Power	222,181	232,500
TOTAL POWER, kWe	**652,155**	**663,400**
AUXILIARY LOAD SUMMARY, kWe		
Coal Handling	510	510
Coal Milling	2,730	2,730
Slag Handling	580	580
WTA Coal Dryer Compressor	9,370	9,370
WTA Coal Dryer Auxiliaries	620	620
Natural Gas Compressors		
Gasifier Steam Generator Circ. Pumps		
Air Separation Unit Auxiliaries	1,003	1,000
Air Separation Unit Main Air Compressor	63,719	63,550
Oxygen Compressor	8,830	8,830
Nitrogen Compressors	33,340	33,340
CO_2 Compressor	31,544	31,560
Boiler Feedwater Pumps	3,851	3,260
Condensate Pump	194	230
Quench Water Pump	760	760
Syngas Recycle Compressor	820	820
Circulating Water Pump	2,931	2,730
Ground Water Pumps	320	310
Cooling Tower Fans	1,911	1,780
Air Cooled Condenser Fans	2,771	2,960
Scrubber Pumps	20	20
Acid Gas Removal	18,390	18,400
Gas Turbine Auxiliaries	998	1,000
Steam Turbine Auxiliaries	96	100
Claus Plant/TGTU Auxiliaries	249	250
Claus Plant TG Recycle Compressor	1,517	1,530
Miscellaneous Balance of Plant	3,000	3,000
Transformer Losses	2,507	2,550
TOTAL AUXILIARIES, kWe	**192,581**	**191,790**
NET POWER, kWe	**459,574**	**471,610**
Net Plant Efficiency, % (HHV)	31.2%	32.1%
Net Plant Heat Rate, Btu/kWh	10,919	10,641
CONDENSER COOLING DUTY, MMBtu/hr	**1,170**	**1,170**
CONSUMABLES		
As-Received Coal Feed, lb/hr	585,971	585,970
Thermal Input, kWt	1,470,705	1,470,704
Raw Water Withdrawal, gpm	3,520	3,404
Raw Water Consumption, gpm	2,842	2,767

3.4. Capital Cost

3.4.1. Total Plant Cost

Table 3-3 shows the total plant cost (TPC) summary for the Reference case and the Reference Bench Mark case.

Table 3-3
Shell Reference Case – Total Plant Cost Summary

Total Plant Cost (June 2011)		Shell Reference Case Bench Mark (Case 2a)	Shell Reference Case (S1B)
Acct. No.	Item/Description	$MM	$MM
1	COAL & SORBENT HANDLING	49.4	49.3
2	COAL & SORBENT PREP & FEED	237.8	237.2
3	FEEDWATER & MISC BOP SYSTEMS	34.4	35.1
4	GASIFIER & ACCESSORIES	751.4	730.6
5A	GAS CLEANUP & PIPING	289.9	287.9
5B.2	CO2 Compression & Drying	66.3	65.7
6	COMBUSTION TURBINE/ACCESSORIES	159.4	159.0
7	HRSG, DUCTING & STACK	54.0	53.9
8	STEAM TURBINE GENERATOR	122.5	126.8
9	COOLING WATER SYSTEM	27.0	28.7
10	ASH/SPENT SORBENT HANDLING SYS	44.4	44.3
11	ACCESSORY ELECTRIC PLANT	105.0	105.0
12	INSTRUMENTATION & CONTROL	32.0	31.9
13	IMPROVEMENTS TO SITE	22.5	22.5
14	BUILDINGS & STRUCTURES	20.9	20.8
	CALCULATED TOTAL COST	2,016.9	1,998.7

3.5. Operating Costs

Table 3-4 shows the operating costs breakdown for the Reference case and the Reference Benchmark case.

Table 3-4
Shell Reference Case – Operating Cost Breakdown

OPERATING COSTS, 2011 $MM/yr	Shell Reference Case Bench Mark (Case 2a)	Shell Reference Case (S1B)
FIXED OPERATING COSTS		
Annual Operting Labor Cost	$7.2	$7.2
Maintainence Labor Cost	$19.5	$19.3
Administration & Support Labor	$6.7	$6.6
Property Taxes and Insurance	$40.3	$40.0
TOTAL FIXED OPERATING COSTS	$73.7	$73.2
VRIABLE OPERATING COSTS (@100% CF)		
NON-FUEL VARIABLE OPERATING COSTS		
Maintenance Material Cost	$45.3	$44.9
Water	$1.5	$1.5
Chemicals		
MU & WT Chemicals	$1.5	$1.4
Other Chemicals & Catalysts	$2.6	$2.6
Waste Disposal	$5.4	$5.4
TOTAL NON_FUEL VARIABLE OPERATING COSTS	$56.3	$55.9
FUEL (@100% CF)	$50.4	$50.4
TOTAL VARIABLE OPERATING COSTS	$106.7	$106.2

3.6. Cost of Electricity

Table 3-3 shows a summary of the power output, CAPEX, OPEX, COE and cost of CO_2 capture for the Reference case and the Reference Bench Mark case.

Table 3-5
Shell Reference Case – Plant Performance and Economic Summary

	Shell Reference Case Bench Mark (Case 2a)	Shell Reference Case (S1B)
CAPEX, $MM		
Total Installed Cost (TIC)	$1,512	1,501
Total Plant Cost (TPC)	$2,017	1,999
Total Overnight Cost (TOC)	$2,472	2,450
OPEX, $MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC_{Fix})	$74	$73
Variable Operating Cost Less Fuel (OC_{VAR})	$56	$56
Fuel Cost (OC_{Fuel})	$50	$50
Power Production, Mwe		
Gas Turbine	430.0	430.9
Steam Turbine	222.2	232.5
Auxiliary Power Consumption	192.6	191.8
Net Power Output	459.6	471.6
Power Generated, MWh/yr (MWH)	4,026,096	4,131,304
COE, excl CO2 TS&M, mills/kWh	144.8	140.0
COE, incl CO2 TS&M, mills/kWh	165.6	156.7
Cost of CO2 Avoided excl CO2 TS&M, $/ton CO2	$79.2	$73.0
Cost of CO2 Avoided incl CO2 TS&M, $/ton CO2	$104.4	$93.1

4. CASE 1 GTI HMB GASIFIER, 100% PRB COAL FEED IGCC PLANT WITH CO_2 CAPTURE

4.1. Process Overview

The Case 1 IGCC power plant is similarly configured as the Reference Case except the Shell SCGP gasifier in the Reference Case is replaced by GTI's HMB gasifier. A simplified block flow diagram for Case 1 IGCC plant is shown in Figure 4-1. In the GTI HMB gasifier, coal is gasified in a molten bed of slag to generate syngas. The HMB gasifier does not have the integral gas quench at the gasifier outlet as the Shell SCGP gasifier. A simplified conceptual sketch of the GTI gasifier is shown below.

Case 1 IGCC power plant is a 100% 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 MW each for a total of 430 MW at the Montana site's elevation. It includes a HRSG and steam turbines to recover waste heat from the GT flue gas for power generation. It is designed to capture CO_2 equivalent to 90% of the raw syngas' carbon content using the double-stage Selexol process. The nominal net IGCC power export capacity after accounting for the auxiliary loads which include CO_2 capture and compression is 454 MWe.

In order to achieve the 90% CO_2 removal target and maintain the same syngas Btu/SCF to the GT, the raw syngas must be converted to hydrogen-rich syngas by the water gas shift (WGS) reaction. The shifted hydrogen-rich syngas has a H_2/CO ratio of ~60 compared to the raw syngas H_2/CO of 0.4. Steam for WGS is provided partly by vaporizing quench water during SG cooling and partly by saturating the water scrubber overhead gas. The balance of the WGS steam requirement is provided by steam addition to the WGS feed gas.

The WGS catalyst also hydrolyzes the COS to H_2S for capture in the AGR. The recovered H_2S from the AGR is converted into elemental sulfur in the Claus plant.

The Case 1 IGCC power plant is consisted of the following major blocks:

- Coal Handling
- Coal Prep, Drying & Feed
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- RTI HMB Gasifier System
- Syngas Cooling (Quench, Scrubbing, Steam Generation)
- Gas Cleaning (Filters, WGS, Hg Removal & AGR)
- CO_2 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
- Electrical Distribution

The IGCC plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 80 percent or 7,000 hours/year at full capacity.

4.2. Process Description

The process descriptions for the various Case 1subsystems are identical to those described for the Shell Reference IGCC case in Section 3.2, except for the gasification section, which is described in detail in Section 4.2.1.

4.2.1. GTI HMB Gasification

The GTI Hybrid Molten Bed (HMB) gasification is a dual-fueled process firing coal and natural gas Typically, natural gas is fired under partial oxidation conditions with oxygen into a bed of molten coal slag to produce a hydrogen-rich gas and heat to drive the endothermic gasification of coal that is charged to the molten bed. The HMB gasification process is described in more details in section 2.2.

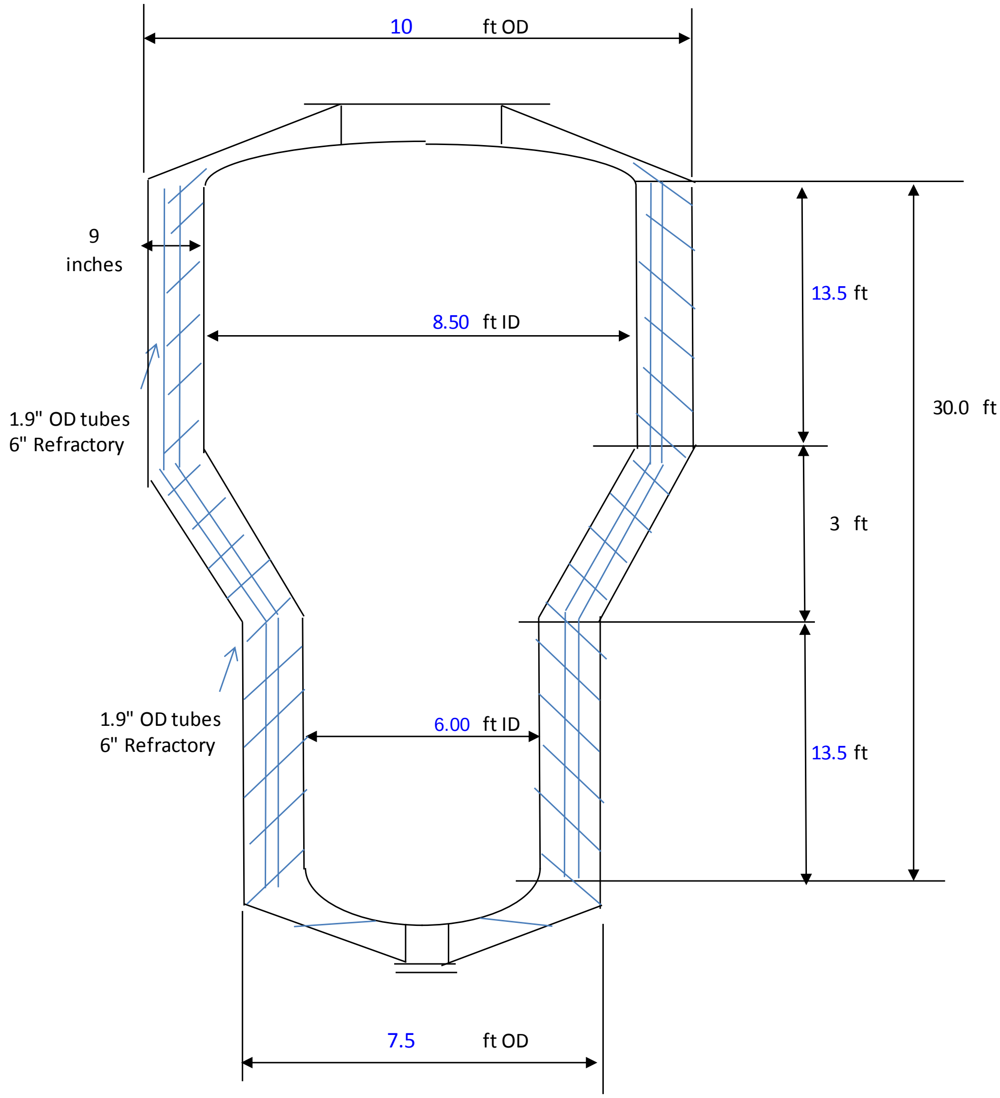
In the Case 1 IGCC power plant, only coal is fired into the molten bed, in a configuration similar to the Shell IGCC Reference Case. GTI has indicated that the molten bed gasifier could run in this 100% coal configuration with no natural gas feed, resulting in a syngas product with a lower H_2/CO ratio.

4.2.2. Gasifier Layout and Dimensions

Figure 4-2 is a conceptual layout of the GTI Hybrid Molten Bed Gasifier based on estimated dimensions provided by GTI. The layout and estimated dimensions formed the basis for the cost estimation for the HMB gasifier.

HMB gasifier tests are currently performed by GTI and the test data will provide refinements to the gasifier dimensions and layout. The final technical details will be provided by GTI in a separate report.

Figure 4-1
GTI HMB Gasifier Conceptual Layout



This conceptual layout depicts the gasifier walls with built in tube banks with a thin layer of castable refractory on the inside. A thin layer of frozen slag forms on the walls, protecting them from abrasion, a process demonstrated with many mineral melts in GTI submerged combustion melters.

Table 4-1 summarizes the dimensions for the GTI MB gasifier.

The cost of the GTI HMB gasifier for the GTI Case 1 is based on the gasifier sizes as shown in the conceptual layout in Figure 4-2. Allowance for burners, steam drums and circulating pumps, and slag removal are included in the gasifier cost.

Table 4-1
Case 1 GTI HMB Gasifier Overall Dimensions

No. of Trains	2
No. of Gasifiers per Train	1
Gasifier Diameter (ID), ft (top)	8.5
Gasifier Diameter (ID),ft (bottom)	6
GTI Gasifier Height, ft (top)	13.5
GTI Gasifier Height, ft (bottom)	13.5
Refractory Thickness, inches, (top)	6
Refractory Thickness, inches, (bottom)	6
Steam Tube OD, inches	1.9
Gasifier Overall Dimensions:	
Shell Diameter (OD), ft (top)	10
Shell Diameter (OD), ft (bottom)	7.5
Swage Height, ft	3
Total Height, ft	30

4.3. Case 1 Performance

Table 4-2 shows the power production and auxiliary load breakdown of the Case 1GTI HMB gasification-based IGCC running on 100% coal feed.

Table 4-2
Case 1 Power Generation and Auxiliary Load Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	**Case 1**
Gas Turbine Power	432,063
Steam Turbine Power	211,142
TOTAL POWER, kWe	**643,205**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	505
Coal Milling	2,704
Slag Handling	483
WTA Coal Dryer Compressor	9,281
WTA Coal Dryer Auxiliaries	614
Natural Gas Compressors	
Gasifier Steam Generator Circ. Pumps	196
Air Separation Unit Auxiliaries	974
Air Separation Unit Main Air Compressor	61,908
Oxygen Compressor	9,336
Nitrogen Compressors	31,572
CO_2 Compressor	31,173
Boiler Feedwater Pumps	3,593
Condensate Pump	247
Quench Water Pump	760
Syngas Recycle Compressor	0
Circulating Water Pump	2,849
Ground Water Pumps	371
Cooling Tower Fans	1,858
Air Cooled Condenser Fans	2,505
Scrubber Pumps	20
Acid Gas Removal	18,199
Gas Turbine Auxiliaries	1,003
Steam Turbine Auxiliaries	91
Claus Plant/TGTU Auxiliaries	247
Claus Plant TG Recycle Compressor	2,770
Miscellaneous Balance of Plant	2,972
Transformer Losses	2,472
TOTAL AUXILIARIES, kWe	**188,704**
NET POWER, kWe	**454,501**
Net Plant Efficiency, % (HHV)	31.2%
Net Plant Heat Rate, Btu/kWh	10,934
CONDENSER COOLING DUTY, MMBtu/hr	**1,058**
CONSUMABLES	
As-Received Coal Feed, lb/hr	580,414
Thermal Input, kWt	1,456,405
Raw Water Withdrawal, gpm	4,744
Raw Water Consumption, gpm	4,074

4.4. Elemental Balance

Tables 4-3 and 4-4 show, respectively, the carbon and sulfur balances for the Case 1 GTI HMB gasifier-based IGCC.

Table 4-3
Case 1 Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
C in Coal Feed	290,603	
C in Natural Gas Feed	-	
C in ASU Air	191	
C in Air to Gas Turbine	797	
C in ASU Vent		191
C in Sour Water		-
C in Slag		-
C in Flyash		1,453
C in Sulfur Product		-
C in Stack Gas		30,350
C in CO2 Product		259,601
Convergence Tolerance		(4)
Total	**291,591**	**291,591**

Table 4-4
Case 1 Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
S in Coal Feed	4,222	
S in Natural Gas Feed	-	
S in ASU Air	-	
S in Air to Gas Turbine	-	
S in ASU Vent		-
S in Sour Water		-
S in Slag		-
S in Flyash		-
Sulfur Product		4,206
Stack Gas		16
S in CO2 Product		-
Convergence Tolerance		0
Total	**4,222**	**4,222**

4.5. Water balance

Water makeup and consumptions are included in the overall utility summary in section 4.8.

4.6. Equipment

The major equipment lists for Case 1 is shown in Table 4-5.

Table 4-5
Case 1 Major Equipment List

VESSELS & TANKS:

Plt No.	Item No.	Item Name	Type	Design Conditions PSIG	Design Conditions deg F	Material of Construction	Quantity per Lot	Units	Inside Diameter Ft	Ht or Tan/Tan Length Ft	Width Ft	Length Ft	Number of Lots	Total Equip Cost $1000
	400C-100	SG Scrubber	Vert	650	450	304Clad shell	2	Vessel	11.5	32.5			2	
	500C-100	Cooled SG KO Drum	Vert	550	450	304Clad shell	2	Vessel	10.0	9.0			2	
	900C-100	SG Water Quench Drum	Vert	650	450	304Clad shell	2	Vessel	12.0	34.0			2	
	600C-100	Cooled Hydrogenated Tail Gas KO	Vert	15	450	304Clad shell	2	Vessel	4.0	8.0			2	
	600C-101	Hydrogenated Tail Gas Compresso	Vert	15	450	304Clad shell	2	Vessel	3.0	7.5			2	

SHELL & TUBE EXCHANGERS AND AIR COOLERS:

Plt No.	Item No.	Item Name	Type	Design PSIG Shell	Design PSIG Tube	Des Temp, deg F Shell	Des Temp, deg F Tube	Material Of Construction Shell	Material Of Construction Tube	Duty MMBtu/Hr	Total Bare Tube Area, Ft2	Physical Arrangement In Series	Physical Arrangement In Parallel	Physical Arrangement Total # Req	Total Equip Cost $1000
	200E-101	O2 Compr Instg Cooler	S&T	319	100	375	375	CS	CS	7.7	2,391	1	2	2	
	400E-103	HP Stm Gen/LTS Feed	Kettle	755	2058	972	706	316SS	316SS	74.3	1,722	1	4	4	
	400E-105	LP Stm Gen/LTS Feed	Kettle	100	2058	535	378	CS	CS	19.4	1,452	1	2	2	
	400E-101	SG Scrubber Fd/Btm Exch	S&T	605	635	473	396	CS	CS	10.0	5,124	1	2	2	
	400E-102	HT WGS Feed/LT WGS SG HX	S&T	735	595	570	512	CS	CS	11.7	2,067	1	2	2	
	400E-202	Quench Water/LTS Elluent	Kettle	755	755	535	413	CS	CS	62.2	3,597	1	2	2	
	900E-100	HP Stm/Quenched SG Cooler #1	Kettle	755	2058	759	706	316SS	316SS	16.9	1,108	1	2	2	
	900E-101	IP Stm/Quenched SG Cooler #1	Kettle	755	2058	695	551	CS	CS	46.2	3,020	1	2	2	
	400E-203	CW/LTS Effluent	S&T	543	100	395	375	316SS	CS	140.9	8,285	1	2	2	
	400E-204	Condensate Preheat/LTS Effluent	S&T	635	100	398	375	316SS	316SS	84.2	8,036	1	2	2	
	500E-100	CO2 Compressor 1st Stg Cooler	S&T	275	100	375	375	304SS	CS	10.0	3,459	1	2	2	
	500E-101	CO2 Compressor 2nd Stg Cooler	S&T	550	100	375	375	304SS	CS	10.0	2,291	1	2	2	
	500E-102	CO2 Compressor 3rd Stg Cooler	S&T	1056	100	375	375	304SS	CS	10.0	2,331	1	2	2	
	500E-104	SC CO2 Cooler	S&T	549	100	972	375	CS	CS	10.0	588	1	2	2	
	600E-100	Sulfur Cooler	S&T	18	100	425	155	316SS	CS	0.2	18	1	2	2	
	600E-101	Hydorgenater Tail Gas Cooler	S&T	11	100	625	155	CS	CS	4.6	269	1	2	2	
	600E-102	Hydrogenated Tail Gas Compresso	S&T	70	100	596	155	CS	CS	2.7	161	1	2	2	
	600E-103	Hydrogenated Tail Gas Compresso	S&T	528	100	623	155	CS	CS	2.1	135	1	2	2	
	700E-100	GT Feed Superheater	S&T	735	500	580	460	316SS	316SS	40.1	2,356	1	2	2	
	100E-100	ASU Air Compr 1st Instg Cooler	S&T	319	100	375	375	CS	CS	8.2	1,409	1	2	2	
	100E-101	ASU Air Compr 2nd Instg Cooler	S&T	319	100	375	375	CS	CS	9.4	1,467	1	2	2	
	100E-102	ASU N2 Primary Compr Instg Cool	S&T	319	100	376	375	CS	CS	6.5	940	1	2	2	

COMPRESSORS, BLOWERS & DRIVERS:

Plt No.	Item No.	Item Name	Type	Design Conditions PSIG	Design Conditions deg F	Material Of Construction Wheel or Impel'r	Material Of Construction Casing	Design Capacity Des Flow SCFM	Design Capacity Inlet PSIA	Design Capacity Delta P PSI	Comp BHP	Driver HP	Driver Type	Total # Req	Total Equip Cost $1000
	200K-100A	NG Compressor 1st Stage	Cent	365	328	CS	CS	-	99.5	250.5	-		Motor	2	
	200K-101A	O2 Compressor 1st Stage	Cent	320.102	327	CS	CS	34,049	124.5	180.6	2,857	3143	Motor	2	
	200K-101B	O2 Compressor 2nd Stage	Cent	754.7	315	CS	CS	34,049	299.602	440.1	2,832	3115	Motor	2	
	600K-100	Hydrogenator TG Compressor 1st	Cent	505	992	CS	CS	3,064	10.1	479.9	1,688	1857	Motor	2	
	600K-101	Hydrogenator TG Compressor 2nd	Cent	505	523	CS	CS	2,760	65	425.0	655	721	Motor	2	
	100K-100A	1st Stg ASU Air Compressor	Cent	52	302	CS	CS	83,891	12.9	24.1	7,753	8528	Motor	4	
	100K-100B	2nd Stg ASU Air Compressor	Cent	95	326	CS	CS	75,313	31.5	48.5	6,503	7154	Motor	4	
	100K-100C	3rd Stg ASU Air Compressor	Cent	205	324	CS	CS	75,313	74.5	115.5	6,526	7178	Motor	4	
	500K-100A	1st Stg CO2 Compressor	Cent	265	188	CS	CS	34,459	148.7	101.3	1,485	1633	Motor	4	
	500K-100B	2nd Stg CO2 Compressor	Cent	515	242	CS	CS	46,819	234.5	265.5	3,109	3420	Motor	4	
	500K-100C	3rd Stg CO2 Compressor	Cent	1020	239	CS	CS	46,819	494.5	510.5	2,907	3197	Motor	4	
	100K-101A	1st Stg Primary N2 Compressor	Cent	163	351	CS	CS	53,428	55.9	92.1	4,942	5436	Motor	4	
	100K-101B	2nd Stg Primary N2 Compressor	Cent	400	341	CS	CS	53,428	142.5	242.5	4,965	5462	Motor	4	
	100K-102	Secondary N2 Compressor	Cent	400	284	CS	CS	15,591	181.5	203.5	1,058	1164	Motor	2	

PUMPS & DRIVERS:

Plt No.	Item No.	Item Name	Type	Design Conditi PSIG	Design Conditi deg F	Material Of Construction Wheel or Impel'r	Material Of Construction Casing	Design Capacity Des Flow GPM	Design Capacity Inlet PSIG	Design Capacity Delta P PSI	Pump BHP	Driver HP	Driver Type	Total # Req	Total Equip Cost $1000
	300G-100	SG Steam Gen BFW Pump	Cent.	790	450	304SS	304SS	510	591	149	56	62	Motor	4	
	400G-100	SG Scrubber Recirc Pump	Cent.	630	450	304SS	304SS	730	560	21	11	12	Motor	4	
	500G-100	SC CO2 Pump	Cent.	2317	450	304SS	304SS	1318	994	1212	1164	1294	Motor	4	

PACKAGED & MISC EQUIPMENT:

Plt No.	Item No.	Item Name	Type	Design Conditions PSIG	Design Conditions deg F	Mat Of Construct	Design Capacity	Remarks	Total # Req	Total Equip Cost $1000
		Particulate Filters				304SS		Capacity Factored		

4.7. Utilities

Table 4-6 shows the utilities summary of Case 1

Table 4-6
Case 1 Utilities Summary

Item Name	Elect. Power	Steam							Water Requirement, 1000 Lbs/Hr													
		1000 Lbs/Hr							Cold Cond	Return Cond	BFW	LP Cond	MP Cond 1	MP Cond 2	Process Effluent	SWS Stripped Water	Open	Open	Raw Makeup	Blowdown	Wastewater to Treatment	Cooling Water	
	KW	SH Process Stm: 715 PSIA/ 1200F	SHHP: 1815 PSIA/ 1000F	HP: 1875 PSIA/ 627F Sat	Process Stm: 720 PSIA / 505 F Sat	IP: 525 PSIA / 472 F Sat	250 PSIA / 786F	LP: 65 PSIA / 298 F Sat	Cold Cond / 90F	Return Cond / 235F	1975 PSIA / 288F	LP Cond/ 293F	MP Cond/ 471F	250 PSIA / 401F					Raw Makeup Water	Blowdown	Waste W / @ 100 F	CW, MMbtu/hr	C.W. circ. GPM
PROCESS/GASIFICATION ISLAND																						
COAL/SLAG HANDLING & MILLING:																						
COAL HANDLING	505	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
COAL MILLING	2,704	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SLAG HANDLING	483	-	-	-	-	-	-	-	-	-	-	-	-	-	-	53	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
WTA DRYING:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
WTA COAL DRYER COMPRESSOR	9,281	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
WTA COAL DRYER AUXILIARIES	614	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	16	1,583
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
AIR SEPARATION UNIT:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ASU AUXILIARIES	974	-	-	-	-	-	29	48	-	-	-	(48)	(29)	-	-	-	-	-	-	-	-	-	-
ASU MAIN COMPRESSOR	61,908	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ASU INTERCOOLER (incl O2 & N2 Intercooling)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	201	20,048
OXYGEN COMPRESSOR	9,336	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	15	1,543
NITROGEN COMPRESSOR	31,572	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NITROGEN BOOST COMPRESSOR	880	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM REFORMER:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NATURAL GAS COMPRESSOR		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NG COMPRESSOR INTERCOOLERS		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
REFORMING STEAM INJECTION		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GASIFIER & SYNGAS COOLING:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GTI GASIFIER WALL STEAM GENERATORS		-	-	-	(105)	-	-	-	-	-	105	-	-	-	-	-	-	-	-	-	-	-	-
SG / PROCESS STEAM GENERATOR		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SG / PROCESS STEAM SUPERHEATER		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SG / HP STEAM GENERATOR		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SG / HP STEAM SPERHEATER		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GASIFIER STM GEN CIRC PUMP	196	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ASH COOLING		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	29	2,857
QUENCH COOLER		-	-	-	-	-	-	-	507	-	-	-	-	-	507	-	-	-	-	-	-	-	-
QUENCH WATER PUMP	760	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS RECYCLE COMPRESSOR		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
QUENCHED SYNGAS COOLER/STEAM GEN		-	-	(38)	-	(97)	-	-	-	-	135	-	-	-	-	-	-	-	-	(1)	-	-	-
IP STEAM LETDOWN		-	-	-	-	24	-	(24)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SCRUBBER, SHIFT & SYNGAS COOLING:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS SCRUBBER		-	-	-	-	-	-	-	90	-	-	-	-	-	(122)	-	-	-	-	-	-	-	-
SCRUBBER PUMPS	20	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SHIFT REACTOR		-	-	176	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
HT SHIFT SYNGAS COOLER/STEAM GEN		-	-	(334)	-	-	-	(42)	-	-	377	-	-	-	-	-	-	-	-	(2)	-	-	-
LT SHIFT SYNGAS COOLER/STEAM GEN		-	-	-	-	-	-	-	1,155	(1,155)	-	-	-	-	-	-	-	-	-	-	-	-	-
LT SHIFT SYNGAS CW COOLER		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	282	28,139
LT SHIFT SYNGAS KO DRUM		-	-	-	-	-	-	-	-	-	-	-	-	-	(319)	-	-	-	-	-	-	-	-
SOUR WATER STRIPPER		-	-	-	-	-	-	31	-	-	-	(31)	-	-	(59)	(388)	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ACID GAS REMOVAL:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
DOUBLE STAGE SELEXOL	18,199	-	-	-	-	-	-	143	-	-	-	(143)	-	-	-	-	-	-	-	-	-	-	-
DOUBLE STAGE SELEXOL COOLING		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	61	6,047
CLAUS PLANT/TGTU AUXILIARIES	247	-	-	-	-	(7)	-	(4)	-	-	16	-	(5)	-	-	-	-	-	-	-	-	-	-
CLAUS PLANT TG RECYCLE COMPRESSOR	1,726	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	19	1,877
CLAUS PLANT TG RECYCLE COMPRESSOR INSTG KO DRUMS		-	-	-	-	-	-	-	-	-	-	-	-	-	(6)	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2 COMPRESSION:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2 COMPRESSORS	31,173	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2 COMPRESSION INTERCOOLER		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	144	14,338
CO2 COMPRESSION KO DRUMS		-	-	-	-	-	-	-	-	-	-	-	-	-	(1)	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS REHEAT:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS REHEATER		-	-	-	-	38	-	58	-	-	-	(58)	(38)	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MISCELLANEOUS		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
BALANCE OF PLANT	3,000	-	-	-	-	-	-	4	-	-	-	(4)	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SUBTOTAL PROCESS/GASIFICATION ISLAND	**173,580**	**-**	**-**	**(196)**	**(105)**	**(42)**	**29**	**213**	**1,752**	**(1,155)**	**634**	**(283)**	**(72)**	**-**	**(0)**	**(335)**	**-**	**-**	**-**	**(3)**	**-**	**765**	**76,432**
POWER ISLAND & STEAM CYCLE																							
GAS TURBINE:																							
GAS TURBINE GENERATOR	(432,063)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GAS TURBINE AUXILIARIES	(1,003)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM TURBINE:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM TURBINE GENERATOR	(211,142)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM TURBINE AUXILIARIES	(91)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MISCELLANEOUS:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TRANSFORMER LOSSES	2,472	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM CYCLE:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM / CONDENSATE IMPORT		-	-	196	105	80	-	46	(1,752)	1,155	-	283	72	-	-	-	-	-	777	3	-	-	-
STEAM / CONDENSATE (EXPORT)		-	-	-	-	(38)	(29)	(259)	-	-	(634)	-	-	-	-	-	-	-	-	-	5	-	-
AIR COOLED CONDENSER FANS	2,505	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CW COOLED SURFACE CONDENSER		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	529	52,845
CONDENSATE PUMPS	247	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
BOILER FEED WATER PUMPS	3,593	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SUBTOTAL POWER ISLAND & STEAM CYCLE	**(635,482)**	**-**	**-**	**196**	**105**	**42**	**(29)**	**(213)**	**(1,752)**	**1,155**	**(634)**	**283**	**72**	**-**	**-**	**-**	**-**	**-**	**777**	**3**	**5**	**529**	**52,845**
COOLING WATER & COOLING TOWER																							
COOLING TOWER:																							
COOLING WATER PRODUCTION		-														335.0			1,260.2		(361.6)	(1,294)	(129,277)
COOLING TOWER FANS	3,126	-																					
GROUND WATER PUMPS	371	-																					
CIRCULATING WATER PUMP	2,038	-																					
SUBTOTAL CW & CT	**5,534**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**335**	**-**	**-**	**1,260**	**-**	**(362)**	**(1,294)**	**(129,277)**
GRAND TOTAL IGCC	**(456,368)**	**-**	**-**	**(0)**	**-**	**0**	**0**	**0**	**-**	**-**	**(0)**	**0**	**(0)**	**-**	**(0)**	**-**	**-**	**-**	**2,037**	**-**	**(356)**	**-**	**-**

4.8. Capital Cost

4.8.1. Total Plant Cost

Table 4-7 shows the total plant cost (TPC) summary of Case 1 compared to the Reference Bench Mark case.

Table 4-7
Case 1 Total Plant Cost Summary

Total Plant Cost (June 2011)		GTI MB Case 1	Shell Reference Bench Mark (Case 2a)
Acct. No.	Item/Description	$MM	$MM
1	COAL & SORBENT HANDLING	49.1	49.4
2	COAL & SORBENT PREP & FEED	236.3	237.8
3	FEEDWATER & MISC BOP SYSTEMS	29.9	34.4
4	GASIFIER & ACCESSORIES	401.5	751.4
5	GAS CLEANUP & PIPING	289.4	289.9
5B.2	CO2 Compression & Drying	65.6	66.3
6	COMBUSTION TURBINE/ACCESSORIES	159.4	159.4
7	HRSG, DUCTING & STACK	53.5	54.0
8	STEAM TURBINE GENERATOR	112.1	122.5
9	COOLING WATER SYSTEM	27.9	27.0
10	ASH/SPENT SORBENT HANDLING SYS	39.6	44.4
11	ACCESSORY ELECTRIC PLANT	104.1	105.0
12	INSTRUMENTATION & CONTROL	31.9	32.0
13	IMPROVEMENTS TO SITE	22.2	22.5
14	BUILDINGS & STRUCTURES	20.9	20.9
	CALCULATED TOTAL COST, $1000	1,643.2	2,016.9

4.9. Operating Costs

Table 4-8 summarizes the operating costs for Case1 compared to the Reference Bench Mark case.

Table 4-8
Case 1 Operating Cost Breakdown

OPERATING COSTS, 2011 $MM/yr	GTI MB Case 1	Shell Reference Case Bench Mark (Case 2a)
FIXED OPERATING COSTS		
Annual Operting Labor Cost	$7.2	$7.2
Maintainence Labor Cost	$15.9	$19.5
Administration & Support Labor	$5.8	$6.7
Property Taxes and Insurance	$32.9	$40.3
TOTAL FIXED OPERATING COSTS	$61.8	$73.7
VRIABLE OPERATING COSTS (@100% CF)		
NON-FUEL VARIABLE OPERATING COSTS		
Maintenance Material Cost	$36.9	$45.3
Water	$2.1	$1.5
Chemicals		
MU & WT Chemicals	$2.0	$1.5
Other Chemicals & Catalysts	$2.6	$2.6
Waste Disposal	$5.4	$5.4
TOTAL NON_FUEL VARIABLE OPERATING COSTS	$49.0	$56.3
FUEL (@100% CF)	$49.9	$50.4
TOTAL VARIABLE OPERATING COSTS	$98.9	$106.7

4.10. Cost of Electricity

Table 4-9 shows a summary of the power output, CAPEX, OPEX, COE and cost of CO_2 capture for Case1 compared to the Reference Bench Mark case.

Table 4-9
Case 1 Plant Performance and Economic Summary

	GTI MB Case 1	Shell Reference Case Bench Mark (Case 2a)
CAPEX, $MM		
Total Installed Cost (TIC)	$1,242	$1,512
Total Plant Cost (TPC)	$1,643	$2,017
Total Overnight Cost (TOC)	$2,012	$2,472
OPEX, $MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC_{Fix})	$62	$74
Variable Operating Cost Less Fuel (OC_{VAR})	$49	$56
Fuel Cost (OC_{Fuel})	$50	$50
Power Production, Mwe		
Gas Turbine	432.1	430.0
Steam Turbine	211.1	222.2
Auxiliary Power Consumption	188.7	192.6
Net Power Output	454.5	459.6
Power Generated, MWh/yr (MWH)	3,981,429	4,026,096
COE, excl CO2 TS&M, mills/kWh	122.8	144.8
COE, incl CO2 TS&M, mills/kWh	143.7	165.6
Cost of CO2 Avoided excl CO2 TS&M, $/ton CO2	$52.6	$79.2
Cost of CO2 Avoided incl CO2 TS&M, $/ton CO2	$78.0	$104.4

.

5. CASE 2: GTI HMB GASIFIER, 55% PRB COAL / 45% NG FEED IGCC PLANT WITH CO_2 CAPTURE

5.1. Process Overview

Case 2 IGCC power plant is configured to use the GTI Hybrid Molten Bed (HMB) Gasifier designed for 55% Montana PRB coal/45% natural gas co-feed. It is designed to generate enough hydrogen-rich fuel gas to fill two advanced GE 7F-turbines rated nominally at 215 MW each for a total of 430 MW at the Montana site's elevation. It includes a HRSG and steam turbines to recover waste heat from the GT flue gas to maximize power generation. It is designed to capture CO_2 equivalent to 90% of the raw syngas' carbon content using the double-stage Selexol process. The nominal net IGCC power export capacity after accounting for the auxiliary loads which include CO_2 capture and compression is 510 MWe. A simplified block flow diagram for Case 2 IGCC plant is shown in Figure 2-3.

Case 2 syngas cooling/heat integration is optimized for high temperature natural gas and steam feed preheat and high pressure steam generation and superheat. With the high temperature feed preheat requirements it was deemed advantages to maximize high level syngas heat recovery instead of using water quench for syngas cooling with the goal to improve the overall plant efficiency. In this heat integration scheme, hot syngas at 2600°F exiting the HMB gasifier is first quenched with cold recycled syngas to below the PRB coal ash fusion temperature of 2,238°F to prevent the deposit of molten ash in the downstream equipment. The quenched syngas at ~2,100°F provides the required duties and temperature driving force for preheating natural gas feed to 900°F, gasifier steam to 1200°F and generating superheated (1,000°F) high pressure steam for the steam turbines.

Case 2 HMB with the Coal/NG co-feed provides the following advantages/disadvantages:

<u>Advantages</u>

- There is less carbon in the feed for the same MMBtu (HHV) of feed.
  - 4.9 mol carbon/MMBtu (HHV) PRB Coal,
  - 2.7 mol carbon/MMBtu (HHV) NG

  For the 55% Coal/45% NG feed, the equivalent carbon content to be processed in the AGR, CO_2 dehydration and compression sections is 20% lower.

- The gasification process is made more efficient than other gasifiers by recuperating heat from its walls and from the hot, raw syngas through endothermic steam reforming of natural gas, enabling chemical energy to be returned as fuel to the gasifier and heat recycle to the gasifier through natural gas and steam preheating.

- The addition of steam/NG feed to the gasifier results in higher gasifier syngas H_2/CO ratio (1.3) due to reforming of natural gas in the gasifier. This reduces the size of the WGS and shift steam requirement.

Disadvantages

- Lower cold gas efficiency (CGE) compared to Case 1

  CGE is defined as $(H_2+CO)_{HHV}$ / $(Feed)_{HHV}$ *100

  The lower CGE for Case 2 is primarily due to the additional coal/NG required to heat the unreacted steam (15,111 lbmol/h) from 1,200°F to 2,600°F to produce the same $(H_2+CO)_{HHV}$ in the syngas as Case 1. Therefore, on the same $(H_2+CO)_{HHV}$ basis, Case 2 has a higher $(Feed)_{HHV}$ and hence a lower CGE.

- More costly high temperature exchangers due to high cost alloy material of construction.

The Case 2 IGCC power plant is consisted of the following major blocks:

- Coal Handling
- Coal Prep, Drying & Feed
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- GTI HMB Gasifier System
- Syngas Cooling (Gas Quench, Scrubbing, Steam Generation)
- Gas Cleaning (Filters, WGS, Hg Removal & AGR)
- CO_2 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
- Electrical Distribution

The IGCC plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 80 percent or 7,000 hours/year at full capacity.

5.2. Process Description

The process descriptions for the various Case 2 subsystems are identical to those described for the Shell Reference IGCC case in Section 3.2, except for the dual-fueled gasification section, which runs on a combination of natural gas and coal, and is described in detail in Section 5.2.1, as well as the syngas cooling section, which is described in Section 5.2.3.

5.2.1. GTI HMB Gasification

The GTI Hybrid Molten Bed (HMB) gasifier gasification is a dual-fueled process firing coal and natural gas . Typically, natural gas is fired under partial oxidation conditions with oxygen into a bed of molten coal slag to produce a hydrogen-rich gas and heat to drive the endothermic gasification of coal that is charged to the molten bed. The HMB gasification process is described in more details in section 2.2.

5.2.2. Gasifier Layout and Dimensions

Figure 5-2 is a conceptual layout of the GTI Hybrid Molten Bed Gasifier based on estimated dimensions provided by GTI. The layout and estimated dimensions formed the basis for the cost estimation for the HMB gasifier.

HMB gasifier tests are currently performed by GTI and the test data will provide refinements to the gasifier dimensions and layout. The final technical details will be provided by GTI in a separate report.

Figure 5-1
GTI HMB Gasifier Conceptual Layout

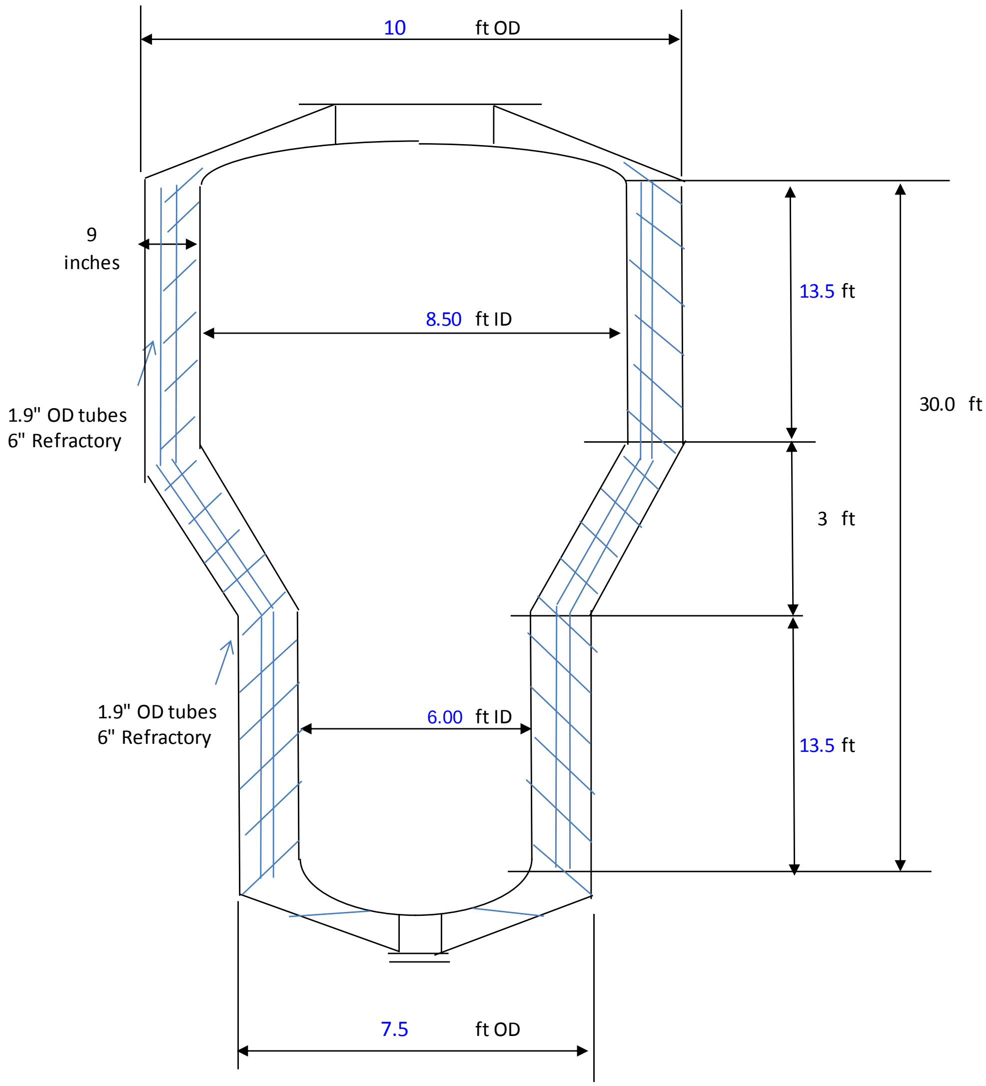


Table 5-1 summarizes the dimensions for the GTI MB gasifier.

Table 5-1
GTI MB Gasifier Overall Dimensions

No. of Trains	2
No. of Gasifiers per Train	1
Gasifier Diameter (ID), ft (top)	8.5
Gasifier Diameter (ID),ft (bottom)	6
GTI Gasifier Height, ft (top)	13.5
GTI Gasifier Height, ft (bottom)	13.5
Refractory Thickness, inches, (bop)	6
Refractory Thickness, inches, (bottom)	6
Steam Tube OD, inches	1.9
Gasifier Overall Dimensions:	
Shell Diameter (OD), ft (top)	10
Shell Diameter (OD), ft (bottom)	7.5
Swage Height, ft	3
Total Height, ft	30

5.2.3. Syngas Cooling & Particulate Filters

Case 2 syngas cooling / heat integration is optimized for natural gas and steam feed preheat and high pressure steam generation and superheat. With the high temperature feed preheat requirements it was deemed advantageous to maximize high level syngas heat recovery instead of using water quench for syngas cooling with the goal of improving the overall plant efficiency. In this heat integration scheme, hot syngas at 2,600°F exiting the HMB gasifier is first quenched with cold recycled syngas to below the PRB coal ash fusion temperature of 2,238°F to prevent the deposit of molten ash in the downstream equipment. The quenched syngas at ~2,100°F provides the required duties and temperature driving force for preheating natural gas feed to 900°F, gasifier steam to 1,200°F and generate superheated (1,000°F) high pressure steam for the steam turbines. After feed preheating, the raw syngas is cooled to 685°F by high temperature steam generation before entering the ceramic particulate filters and cyclones. Any remaining particulate matters in the syngas will be removed by these particulate filters and cyclones.

The filtered syngas is then cooled to 450°F through a series of steam generators generating steam for the steam cycle. The cooled syngas enters a water scrubber to remove any chlorides and remaining particulates. The scrubbed syngas is sent to WGS. Part of the scrubber bottoms is used for slag water bath makeup.

5.3. Case 2 Performance

Table 5-2 shows the power production and auxiliary load breakdown of the Case 2 co-fired GTI HMB gasification-based IGCC running on 55% coal feed/45% natural gas feed. Noted that the steam turbine produced 57 MW more power with the steam generated from high temperature syngas cooling when compared to the water quench case (Case 1).

Table 5-2
Case 2 Power Generation and Auxiliary Load Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	GTI Case 2, PRBCC 55%Coal/45% NG
Gas Turbine Power	431,306
Steam Turbine Power	267,585
TOTAL POWER, kWe	**698,890**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	291
Coal Milling	1,557
Slag Handling	278
WTA Coal Dryer Compressor	5,344
WTA Coal Dryer Auxiliaries	354
Natural Gas Compressors	5,658
Gasifier Steam Generator Circ. Pumps	195
Air Separation Unit Auxiliaries	1,051
Air Separation Unit Main Air Compressor	66,805
Oxygen Compressor	10,278
Nitrogen Compressors	32,463
CO_2 Compressor	26,159
Boiler Feedwater Pumps	4,886
Condensate Pump	262
Quench Water Pump	0
Syngas Recycle Compressor	1,116
Circulating Water Pump	3,366
Ground Water Pumps	341
Cooling Tower Fans	2,195
Air Cooled Condenser Fans	3,268
Scrubber Pumps	17
Acid Gas Removal	15,274
Gas Turbine Auxiliaries	1,001
Steam Turbine Auxiliaries	115
Claus Plant/TGTU Auxiliaries	142
Claus Plant TG Recycle Compressor	2,153
Miscellaneous Balance of Plant	1,711
Transformer Losses	2,686
TOTAL AUXILIARIES, kWe	**188,964**
NET POWER, kWe	**509,926**
Net Plant Efficiency, % (HHV)	33.4%
Net Plant Heat Rate, Btu/kWh	10,205
CONDENSER COOLING DUTY, MMBtu/hr	**1,380**
CONSUMABLES	
As-Received Coal Feed, lb/hr	334,168
Thermal Input (Coal + NG) , kWt	1,525,059
Raw Water Withdrawal, gpm	4,270
Raw Water Consumption, gpm	3,740

5.4. Elemental Balance

Tables 5-3 and 5-4 show, respectively, the carbon and sulfur balances for the Case 2 co-fired GTI HMB gasifier-based IGCC.

Table 5-3
Case 2 Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
C in Coal Feed	167,312	
C in Natural Gas Feed	74,888	
C in ASU Air	206	
C in Air to Gas Turbine	797	
C in ASU Vent		206
C in Sour Water		-
C in Slag		-
C in Flyash		-
C in Sulfur Product		-
C in Stack Gas		25,175
C in CO2 Product		217,825
Convergence Tolerance		(2)
Total	**243,204**	**243,204**

Table 5-4
Case 2 Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
S in Coal Feed	2,431	
S in Natural Gas Feed	-	
S in ASU Air	-	
S in Air to Gas Turbine	-	
S in ASU Vent		-
S in Sour Water		-
S in Slag		-
S in Flyash		-
Sulfur Product		2,421
Stack Gas		9
S in CO2 Product		-
Convergence Tolerance		0
Total	**2,431**	**2,431**

5.5. Water balance

Water makeup and consumptions are included in the overall utility summary in section 5.8.

5.6. Equipment

The major equipment lists for Case 2 is shown in Table 5-5.

Table 5-5
Case 2 Major Equipment List

VESSELS & TANKS:

Plt No.	Item No.	Item Name	Type	Design Conditions PSIG	Design Conditions deg F	Material of Construction	Quantity per Lot	Units	Inside Diameter Ft	Ht or Tan/Tan Length Ft	Width Ft	Length Ft	Number of Lots	Total Equip Cost $1000
	400C-100	SG Scrubber	Vert	650	450	304Clad shell	2	Vessel	11.0	32.0			2	
	500C-100	Cooled SG KO Drum	Vert	550	450	304Clad shell	2	Vessel	10.0	8.5			2	
	600C-100	Cooled Hydrogenated Tail Gas KO	Vert	15	450	304Clad shell	2	Vessel	4.0	8.0			2	
	600C-101	Hydrogenated Tail Gas Compresso	Vert	15	450	304Clad shell	2	Vessel	3.0	7.5			2	

SHELL & TUBE EXCHANGERS AND AIR COOLERS:

Plt No.	Item No.	Item Name	Type	Design PSIG Shell	Design PSIG Tube	Des Temp, deg F Shell	Des Temp, deg F Tube	Material of Construction Shell	Material of Construction Tube	Duty MMBtu/Hr	Total Bare Tube Area, Ft2	Physical Arrangement In Series	Physical Arrangement In Parallel	Physical Arrangement Total # Req	Total Equip Cost $1000
	300E-107	Process Stm Gen/Quenched SG	Kettle	640	760	1050	580	316SS	316SS	89.6	2,634	1	2	2	
	200E-100	NG Compr Instg Cooler	S&T	282	100	375	375	CS	CS	4.5	1,728	1	2	2	
	200E-101	O2 Compr Instg Cooler	S&T	323	100	375	375	CS	CS	8.5	2,609	1	2	2	
	300E-100	HP Steam Superheater/HT Quenche	S&T	655	2027	1344	1075	316SS	316SS	65.9	4,355	1	2	2	
	300E-102	Process Steam Superheater/HT Qu	Kettle	640	2027	1275	1275	316SS	316SS	54.6	1,606	1	2	2	
	300E-103	NG Preheater 1	S&T	645	767	1096	725	316SS	316SS	14.6	861	1	2	2	
	300E-104	NG Preheater 2	S&T	650	762	1130	975	316SS	316SS	10.7	887	1	2	2	
	400E-103	HP Stm Gen/LTS Feed	Kettle	640	2027	841	706	316SS	316SS	61.4	2,643	1	2	2	
	400E-105	IP Stm Gen/LTS Feed	Kettle	640	2027	635	550	CS	CS	29.0	2,349	1	2	2	
	400E-101	SG Scrubber Fd/Btm Exch	S&T	605	635	438	375	CS	CS	3.1	298	1	2	2	
	400E-102	HT WGS Feed/LT WGS SG HX	S&T	735	595	535	510	316SS	316SS	17.9	5,839	1	2	2	
	400E-202	LP Stm Gen/LTS Elluent	Kettle	100	2027	541	378	CS	CS	24.0	1,442	1	4	4	
	900E-100	HP Stm/Quenched SG Cooler #1	Kettle	640	2027	760	706	316SS	CS	16.9	1,080	1	2	2	
	900E-101	IP Stm/Quenched SG Cooler #1	Kettle	640	2027	705	551	316SS	CS	48.5	2,666	1	2	2	
	900E-102	LP Stm/Quenched SG Cooler #1	Kettle	640	2027	576	376	CS	CS	4.9	269	1	2	2	
	300E-101	HP Steam Generator/HT Quenched	Kettle	640	2027	705	705	316SS	316SS	234.3	6,890	1	2	2	
	400E-203	CW/LTS Effluent	S&T	543	100	395	375	316SS	CS	67.8	3,985	1	2	2	
	400E-204	Condensate Preheat/LTS Effluent	S&T	635	100	398	375	CS	CS	125.8	12,009	1	2	2	
	500E-100	CO2 Compressor 1st Stg Cooler	S&T	275	100	375	375	CS	CS	3.1	1,072	1	2	2	
	500E-101	CO2 Compressor 2nd Stg Cooler	S&T	550	100	375	375	CS	CS	3.1	713	1	2	2	
	500E-102	CO2 Compressor 3rd Stg Cooler	S&T	1056	100	375	375	CS	CS	3.1	725	1	2	2	
	500E-104	SC CO2 Cooler	S&T	549	100	841	375	CS	CS	3.1	183	1	2	2	
	600E-100	Sulfur Cooler	S&T	18	100	425	155	316SS	CS	0.1	10	1	2	2	
	600E-101	Hydorgenater Tail Gas Cooler	S&T	11	100	625	155	CS	CS	3.3	197	1	2	2	
	600E-102	Hydrogenated Tail Gas Compresso	S&T	70	100	592	155	CS	CS	2.1	125	1	2	2	
	600E-103	Hydrogenated Tail Gas Compresso	S&T	528	100	619	155	CS	CS	1.6	106	1	2	2	
	700E-100	GT Feed Superheater	S&T	735	500	580	460	CS	CS	38.7	2,275	1	2	2	
	100E-100	ASU Air Compr 1st Instg Cooler	S&T	323	100	375	375	CS	CS	8.9	1,521	1	2	2	
	100E-101	ASU Air Compr 2nd Instg Cooler	S&T	323	100	375	375	CS	CS	10.1	1,583	1	2	2	
	100E-102	ASU N2 Primary Compr Instg Cool	S&T	323	100	376	375	CS	CS	7.1	1,014	1	2	2	

COMPRESSORS, BLOWERS & DRIVERS:
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Plt No.	Item No.	Item Name	Type	Design Conditic PSIG	Design Conditic deg F	Material of Construction Wheel or Impel'r	Material of Construction Casing	Design Capacity Des Flow SCFM	Design Capacity Inlet PSIA	Design Capacity Delta P PSI	Design Capacity Comp BHP	Driver HP	Driver Type	Total # Req	Total Equip Cost $1000
	200K-100A	NG Compressor 1st Stage	Cent	365	328	CS	CS	18,923	99.5	250.5	2,202	2422	Motor	2	
	200K-100B	NG Compressor 2nd Stage	Cent	746.7	241	CS	CS	18,923	344.5	387.2	1,246	1370	Motor	2	
	200K-101A	O2 Compressor 1st Stage	Cent	323.16	329	CS	CS	37,010	124.5	183.7	3,146	3460	Motor	2	
	200K-101B	O2 Compressor 2nd Stage	Cent	769.7	318	CS	CS	37,010	302.66	452.0	3,117	3429	Motor	2	
	200K-102	Recycle SG Compressor	Cent	630	527	CS	CS	56,928	569.2	45.8	680	748	Motor	2	
	600K-100	Hydrogenator TG Compressor 1st	Cent	505	992	CS	CS	2,381	10.1	479.9	1,312	1443	Motor	2	
	600K-101	Hydrogenator TG Compressor 2nd	Cent	505	523	CS	CS	2,146	65	425.0	509	560	Motor	2	
	100K-100A	1st Stg ASU Air Compressor	Cent	52	302	CS	CS	90,527	12.9	24.1	8,366	9203	Motor	4	
	100K-100B	2nd Stg ASU Air Compressor	Cent	95	326	CS	CS	81,270	31.5	48.5	7,018	7719	Motor	4	
	100K-100C	3rd Stg ASU Air Compressor	Cent	205	324	CS	CS	81,270	74.5	115.5	7,042	7746	Motor	4	
	500K-100A	1st Stg CO2 Compressor	Cent	265	189	CS	CS	28,946	148.7	101.3	1,248	1373	Motor	4	
	500K-100B	2nd Stg CO2 Compressor	Cent	515	242	CS	CS	39,290	234.5	265.5	2,609	2870	Motor	4	
	500K-100C	3rd Stg CO2 Compressor	Cent	1020	239	CS	CS	39,290	494.5	510.5	2,439	2683	Motor	4	
	100K-101A	1st Stg Primary N2 Compressor	Cent	163	351	CS	CS	57,654	55.9	92.1	5,333	5866	Motor	4	
	100K-101B	2nd Stg Primary N2 Compressor	Cent	400	341	CS	CS	57,654	142.5	242.5	5,358	5894	Motor	4	
	100K-102	Secondary N2 Compressor	Cent	400	284	CS	CS	16,824	181.5	203.5	1,142	1256	Motor	2	

PUMPS & DRIVERS:
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Plt No.	Item No.	Item Name	Type	Design Conditi PSIG	Design Conditi deg F	Material of Construction Wheel or Impel'r	Material of Construction Casing	Design Capacity Des Flow GPM	Design Capacity Inlet PSIG	Design Capacity Delta P PSI	Design Capacity Pump BHP	Driver HP	Driver Type	Total # Req	Total Equip Cost $1000
	300G-100	SG Steam Gen BFW Pump	Cent.	790	450	304SS	304SS	324	590	149	35	39	Motor	4	
	400G-100	SG Scrubber Recirc Pump	Cent.	630	450	304SS	304SS	396	560	21	6	7	Motor	4	
	500G-100	SC CO2 Pump	Cent.	2317	450	304SS	304SS	1103	995	1212	974	1082	Motor	4	

PACKAGED & MISC EQUIPMENT:
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Plt No.	Item No.	Item Name	Type	Design Conditions PSIG	Design Conditions deg F	Mat Of Construct	Design Capacity	Remarks	Total # Req	Total Equip Cost $1000
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		Particulate Filters				304SS		Capacity Factored		

5.7. Utilities

Table 5-6 shows the utilities summary of the Case 2.

Table 5-6
Case 2 Utilities Summary

Item Name	Elect. Power KW	Steam 1000 Lbs/Hr SH Process Stm: 715 PSIA/ 1200F	SHHP: 1815 PSIA/ 1000F	HP: 1875 PSIA/ 627F Sat	Process Stm: 720 PSIA / 505 F Sat	IP: 525 PSIA / 472 F Sat	250 PSIA / 786F	LP: 65 PSIA / 298 F Sat	Water Requirement, 1000 Lbs/Hr Cold Cond: Cold Cond / 90F	Return Cond: Return Cond / 235F	BFW: 1975 PSIA / 288F	LP Cond: LP Cond/ 293F	MP Cond 1: MP Cond/ 471F	MP Cond 2: 250 PSIA / 401F	Process Effluent	SWS Stripped Water	Open	Open	Raw Makeup: Raw Makeup Water	Blowdown: Blowdown	Wastewater to Treatment: Waste W / @ 100 F	Cooling Water CW, MMbtu/hr	Cooling Water C.W. circ. GPM
PROCESS/GASIFICATION ISLAND																							
COAL/SLAG HANDLING & MILLING:																							
COAL HANDLING	291	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
COAL MILLING	1,557	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SLAG HANDLING	278	-	-	-	-	-	-	-	-	-	-	-	-	-	-	31	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
WTA DRYING:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
WTA COAL DRYER COMPRESSOR	5,344	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
WTA COAL DRYER AUXILIARIES	354	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	9	912
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
AIR SEPARATION UNIT:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ASU AUXILIARIES	1,051	-	-	-	-	-	31	51	-	-	-	(51)	(31)	-	-	-	-	-	-	-	-	-	-
ASU MAIN COMPRESSOR	66,805	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ASU INTERCOOLER (incl O2 & N2 Intercooling)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	217	21,634
OXYGEN COMPRESSOR	10,278	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	17	1,691
NITROGEN COMPRESSOR	32,463	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NITROGEN BOOST COMPRESSOR	507	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM REFORMER:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NATURAL GAS COMPRESSOR	5,658	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NG COMPRESSOR INTERCOOLERS	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	9	897
REFORMING STEAM INJECTION	-	282	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GASIFIER & SYNGAS COOLING:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GTI GASIFIER WALL STEAM GENERATORS	-	-	-	-	(104)	-	-	-	-	-	104	-	-	-	-	-	-	-	-	(1)	-	-	-
SG / PROCESS STEAM GENERATOR	-	-	-	-	(178)	-	-	-	-	-	179	-	-	-	-	-	-	-	-	(1)	-	-	-
SG / PROCESS STEAM SUPERHEATER	-	(282)	-	-	282	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SG / HP STEAM GENERATOR	-	-	-	(480)	-	-	-	-	-	-	483	-	-	-	-	-	-	-	-	(2)	-	-	-
SG / HP STEAM SPERHEATER	-	-	(480)	480	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GASIFIER STM GEN CIRC PUMP	195	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ASH COOLING	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	16	1,645
QUENCH COOLER	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
QUENCH WATER PUMP	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS RECYCLE COMPRESSOR	1,116	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
QUENCHED SYNGAS COOLER/STEAM GEN	-	-	-	(38)	-	(102)	-	(10)	-	-	151	-	-	-	-	-	-	-	-	(1)	-	-	-
IP STEAM LETDOWN	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SCRUBBER, SHIFT & SYNGAS COOLING:	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS SCRUBBER	-	-	-	-	-	-	-	-	69	-	-	-	-	-	(70)	-	-	-	-	-	-	-	-
SCRUBBER PUMPS	17	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SHIFT REACTOR	-	-	-	200	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
HT SHIFT SYNGAS COOLER/STEAM GEN	-	-	-	(138)	-	(61)	-	-	-	-	200	-	-	-	-	-	-	-	-	(1)	-	-	-
LT SHIFT SYNGAS COOLER/STEAM GEN	-	-	-	-	-	-	-	(104)	1,730	(1,730)	104	-	-	-	-	-	-	-	-	1	-	-	-
LT SHIFT SYNGAS CW COOLER	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	136	13,537
LT SHIFT SYNGAS KO DRUM	-	-	-	-	-	-	-	-	-	-	-	-	-	-	(256)	-	-	-	-	-	-	-	-
SOUR WATER STRIPPER	-	-	-	-	-	-	-	18	-	-	-	(18)	-	-	332	(295)	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ACID GAS REMOVAL:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
DOUBLE STAGE SELEXOL	15,274	-	-	-	-	-	-	120	-	-	-	(120)	-	-	-	-	-	-	-	-	-	-	-
DOUBLE STAGE SELEXOL COOLING	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	51	5,064
CLAUS PLANT/TGTU AUXILIARIES	142	-	-	-	-	(4)	-	(3)	-	-	9	-	(3)	-	-	-	-	-	-	-	-	-	-
CLAUS PLANT TG RECYCLE COMPRESSOR	1,341	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	14	1,416
CLAUS PLANT TG RECYCLE COMPRESSOR INSTG KO DRUMS	-	-	-	-	-	-	-	-	-	-	-	-	-	-	(4)	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2 COMPRESSION:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2 COMPRESSORS	26,159	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2 COMPRESSION INTERCOOLER	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	120	12,031
CO2 COMPRESSION KO DRUMS	-	-	-	-	-	-	-	-	-	-	-	-	-	-	(1)	-	-	-	-	-	-	-	-
	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS REHEAT:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS REHEATER	-	-	-	-	-	36	-	56	-	-	-	(56)	(36)	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MISCELLANEOUS		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
BALANCE OF PLANT	3,000	-	-	-	-	-	-	4	-	-	-	(4)	-	-	-	-	-	-	-	-	-	-	-
	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SUBTOTAL PROCESS/GASIFICATION ISLAND	**171,829**	**-**	**(480)**	**24**	**-**	**(130)**	**31**	**132**	**1,800**	**(1,730)**	**1,229**	**(249)**	**(70)**	**-**	**(0)**	**(265)**	**-**	**-**	**-**	**(5)**	**-**	**589**	**58,826**
POWER ISLAND & STEAM CYCLE																							
GAS TURBINE:																							
GAS TURBINE GENERATOR	(431,306)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GAS TURBINE AUXILIARIES	(1,001)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM TURBINE:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM TURBINE GENERATOR	(267,585)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM TURBINE AUXILIARIES	(115)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MISCELLANEOUS:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TRANSFORMER LOSSES	2,686	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM CYCLE:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM / CONDENSATE IMPORT		-	480	(24)	-	167	-	116	(1,800)	1,730	-	249	70	-	-	-	-	-	558	5	-	-	-
STEAM / CONDENSATE (EXPORT)		-	-	-	-	(36)	(31)	(249)	-	-	(1,229)	-	-	-	-	-	-	-	-	-	7	-	-
AIR COOLED CONDENSER FANS	3,268	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CW COOLED SURFACE CONDENSER		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	690	68,926
CONDENSATE PUMPS	262	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
BOILER FEED WATER PUMPS	4,886	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SUBTOTAL POWER ISLAND & STEAM CYCLE	**(688,904)**	**-**	**480**	**(24)**	**-**	**130**	**(31)**	**(132)**	**(1,800)**	**1,730**	**(1,229)**	**249**	**70**	**-**	**-**	**-**	**-**	**-**	**558**	**5**	**7**	**690**	**68,926**
COOLING WATER & COOLING TOWER																							
COOLING TOWER:																							
COOLING WATER PRODUCTION		-														335.0			1,260.2		(361.6)	(1,294)	(129,277)
COOLING TOWER FANS	3,089	-																					
GROUND WATER PUMPS	340	-																					
CIRCULATING WATER PUMP	2,014	-																					
SUBTOTAL CW & CT	**5,443**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**335**	**-**	**-**	**1,260**	**-**	**(362)**	**(1,294)**	**(129,277)**
GRAND TOTAL IGCC	**(511,633)**	**-**	**0**	**0**	**-**	**0**	**0**	**(0)**	**-**	**-**	**(0)**	**0**	**0**	**-**	**(0)**	**70**	**-**	**-**	**1,819**	**(0)**	**(355)**	**(15)**	**(1,524)**

5.8. Capital Cost

5.8.1. Total Plant Cost

Table 5-7 shows the total plant cost (TPC) summary of Case 2 compared to the Reference case.

Table 5-7
Case 2 Total Plant Cost Summary

Total Plant Cost (June 2011)		GTI MB Case 2	Shell Reference Case Bench Mark (Case 2a)
Acct. No.	Item/Description	$MM	$MM
1	COAL & SORBENT HANDLING	34.9	49.4
2	COAL & SORBENT PREP & FEED	164.2	237.8
3	FEEDWATER & MISC BOP SYSTEMS	36.1	34.4
4	GASIFIER & ACCESSORIES	432.2	751.4
5	GAS CLEANUP & PIPING	269.7	289.9
5B.2	CO2 Compression & Drying	56.2	66.3
6	COMBUSTION TURBINE/ACCESSORIES	159.4	159.4
7	HRSG, DUCTING & STACK	54.2	54.0
8	STEAM TURBINE GENERATOR	140.6	122.5
9	COOLING WATER SYSTEM	28.3	27.0
10	ASH/SPENT SORBENT HANDLING SYS	28.1	44.4
11	ACCESSORY ELECTRIC PLANT	105.7	105.0
12	INSTRUMENTATION & CONTROL	31.9	32.0
13	IMPROVEMENTS TO SITE	22.1	22.5
14	BUILDINGS & STRUCTURES	20.7	20.9
	CALCULATED TOTAL COST	1,584.2	2,016.9

5.9. Operating Costs

Table 5-8 summarizes the operating costs for GTI HMB Case 2 and the Shell Reference Bench Mark case.

Table 5-8
Case 2 Operating Cost Breakdown

OPERATING COSTS, 2011 $MM/yr	GTI MB Case 2	Shell Reference Case Bench Mark (Case 2a)
FIXED OPERATING COSTS		
Annual Operting Labor Cost	$7.2	$7.2
Maintainence Labor Cost	$15.3	$19.5
Administration & Support Labor	$5.6	$6.7
Property Taxes and Insurance	$31.7	$40.3
TOTAL FIXED OPERATING COSTS	$59.9	$73.7
VRIABLE OPERATING COSTS (@100% CF)		
NON-FUEL VARIABLE OPERATING COSTS		
Maintenance Material Cost	$35.6	$45.3
Water	$1.9	$1.5
Chemicals		
MU & WT Chemicals	$1.8	$1.5
Other Chemicals & Catalysts	$1.9	$2.6
Waste Disposal	$3.0	$5.4
TOTAL NON_FUEL VARIABLE OPERATING COSTS	$44.2	$56.3
FUEL (@100% CF)		
Coal	$28.7	
Natural Gas	$106.2	$50.4
TOTAL VARIABLE OPERATING COSTS	$179.1	$106.7

5.10. Cost of Electricity

Table 5-9 shows a summary of the power output, CAPEX, OPEX, COE and cost of CO_2 capture for Case 2.

Table 5-9
Case 2 Plant Performance and Economic Summary

	GTI MB Case 2	Shell Reference Case Bench Mark (Case 2a)
CAPEX, $MM		
Total Installed Cost (TIC)	$1,198	$1,512
Total Plant Cost (TPC)	$1,584	$2,017
Total Overnight Cost (TOC)	$1,960	$2,472
OPEX, $MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC_{Fix})	$60	$74
Variable Operating Cost Less Fuel (OC_{VAR})	$44	$56
Fuel Cost (OC_{Fuel})	$135	$50
Power Production, Mwe		
Gas Turbine	431.3	430.0
Steam Turbine	267.6	222.2
Auxiliary Power Consumption	189.0	192.6
Net Power Output	509.9	459.6
Power Generated, MWh/yr (MWH)	4,466,954	4,026,096
COE, excl CO2 TS&M, mills/kWh	125.0	144.8
COE, incl CO2 TS&M, mills/kWh	140.7	165.6
Cost of CO2 Avoided excl CO2 TS&M, $/ton CO2	$53.3	$79.2
Cost of CO2 Avoided incl CO2 TS&M, $/ton CO2	$71.6	$104.4

6. CASE 3: GTI HMB GASIFIER/STEAM REFORMER, 55% PRB COAL / 45% NG FEED IGCC PLANT WITH CO_2 CAPTURE

6.1. Process Overview

Case 3 IGCC power plant is configured to use the GTI Hybrid Molten Bed (HMB) Gasifier and an external steam reformer designed to generate enough hydrogen-rich fuel gas to fill two advanced GE 7F-turbines rated nominally at 215 MW each for a total of 430 MW at the Montana site's elevation. Like the Case 2 co-fired GTI HMB gasifier-based IGCC, the Case 3 HMB Gasifier/steam reformer syngas generator is designed for 55% Montana PRB coal/45% natural gas co-feed (HHV basis). Figure 6-1 shows a block flow diagram of Case 3.

The Case 3 IGCC power plant includes a HRSG and steam turbines to recover waste heat from the GT flue gas to maximize power generation. It is designed to capture CO_2 equivalent to 90% of the raw syngas' carbon content using the double-stage Selexol process. The nominal net IGCC power export capacity after accounting for the auxiliary loads is 483 MWe.

Case 3 syngas cooling/heat integration is optimized to provide the steam reforming duty and for preheating the natural gas and steam feed to improve thermal efficiency of the power plant. The remaining syngas cooling duty is available for high pressure steam generation and superheat. In this heat integration scheme, hot syngas at 2,600°F exiting the HMB gasifier is heat exchanged with the steam reformer to provide the reforming duty. The cooled syngas exits the reformer at ~1,800°F to provide the required duties and temperature driving force for preheating natural gas feed and reforming steam to 900°F and 1,200°F respectively, as well as to generate superheated (1,000°F) high pressure steam for the steam turbines.

Case 3 GTI MB gasifier/reformer scheme with the Coal/NG co-feed provides the following advantages/disadvantages:

<u>Advantages</u>

- As in the Case 2 co-fired IGCC, there is less carbon in feed for the same MMBtu (HHV) of feed.
  - 4.9 mol carbon / MMBtu (HHV) PRB Coal,
  - 2.7 mol carbon / MMBtu (HHV) NG

  For the 55% Coal/45% NG feed, the equivalent carbon content to be processed in the AGR, CO_2 dehydration and compression sections is 20% lower.

- The gasification process is made more efficient than other gasifiers by recuperating heat from its walls and from the hot, raw syngas through endothermic steam reforming of natural gas, enabling chemical energy to be returned as fuel to the gasifier and heat recycle to the gasifier through natural gas and steam preheats.

- The steam reforming of natural gas results in higher gasifier syngas H_2/CO ratio (1.4). This reduces the size of the WGS and shift steam requirement.

- Higher cold gas efficiency (CGE) compared to the Case 1 & Case 2 because of the recycle of heat to the steam reformer resulting in lower syngas temperature exiting the gasification system.

Disadvantages

- Uncertainty in steam reformer designed for natural gas + syngas operation
- More costly high temperature exchangers due to high cost alloy material of construction

The Case 3 IGCC power plant is consisted of the following major blocks:

- Coal Handling
- Coal Prep, Drying & Feed
- Feed Water & Miscellaneous BOP Systems
- Air Separation Unit (ASU)
- GTI HMB Gasifier System including Steam Reformer
- Syngas Cooling (Gas Quench, Scrubbing, Steam Generation)
- Gas Cleaning (Filters, WGS, Hg Removal & AGR)
- CO_2 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
- Electrical Distribution

The IGCC plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 80 percent or 7,000 hours/year at full capacity.

6.2. Process Description

The process descriptions for the various Case 3 subsystems are identical to those described for the Shell Reference IGCC case in Section 3.2, except for:

- The gasifier, which fires reformed syngas instead of natural gas, in the presence of oxygen, into the molten slag, as described in Section 6.2.1

- The addition of the steam reformer, as described in Section 6.2.3, which converts the natural gas feed into reformed syngas to be fired into the HMB gasifier.

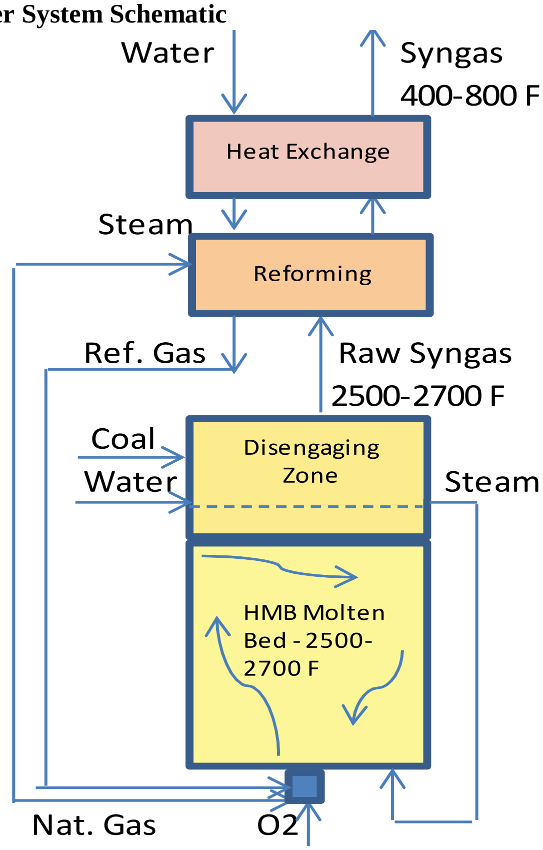
- The syngas cooling section, which is described in Section 6.2.4.

6.2.1. GTI HMB Gasification

In the HMB gasifier for Case 3, in place of natural gas, reformed syngas and oxygen are fired under partial oxidation conditions upward into a bed of molten coal slag. The heat and gases generated drive the gasification process. Evaporative cooling walls generate steam for the external steam reformer to increase process efficiency. Optimization of the coal, natural gas, oxygen, and steam and their ratios to the gasifier/steam reformer generates a syngas with higher H_2/CO ratio of 1.4 to minimize the water gas shift and shift steam requirements. The following is a conceptual description of the HMB gasifier. GTI will provide additional details into the technical development and operational aspects of the HMB gasifier in another report

The HMB gasifier operates at 625 psia and contains a bed of molten slag at 2,600°F maintained by the combustion of reformed syngas and oxygen fired directly into the molten bath. The gasifier walls are built of tube banks with a thin layer of castable refractory on the inside. A thin layer of frozen slag forms on the walls, protecting them from abrasion, a process demonstrated with many mineral melts in GTI submerged combustion melters. Boiler feed water is heated by the wall tube banks to form medium pressure steam. That steam is injected into the steam reformer, providing reactant for steam natural gas reactions while recuperating heat lost from the gasifier walls. Gasifier syngas at 2,600°F provides heat for the steam reformer. The reformer syngas contains H_2, CO and ~10% CH_4 at 1,500°F is charged to the gasifier where the CH_4 is further converted into H_2, CO and CO_2. This innovative method of recovering heat from hot syngas improves overall plant efficiency. A simple schematic of the Case 3 gasifier/steam reformer system is shown in Figure 6-1 below.

Figure 6-1
Case 3 Gasifier/Steam Reformer System Schematic



6.2.2. Gasifier Layout and Dimensions

Figure 6-2 is a conceptual layout of the GTI Hybrid Molten Bed Gasifier based on estimated dimensions provided by GTI. The layout and estimated dimensions formed the basis for the cost estimation for the HMB gasifier.

HMB gasifier tests are currently performed by GTI and the test data will provide refinements to the gasifier dimensions and layout. The final technical details will be provided by GTI in a separate report.

Figure 6-2
GTI HMB Gasifier Conceptual Layout

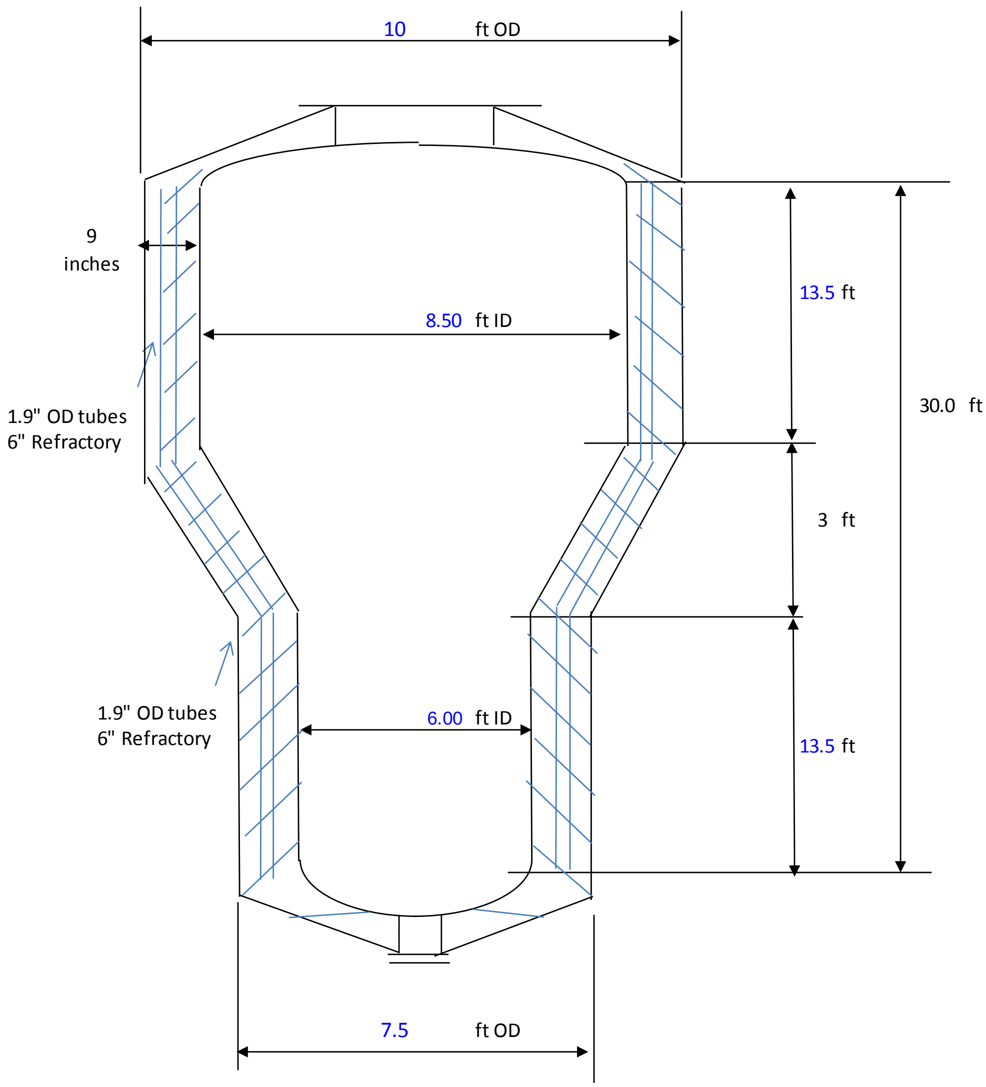


Table 6-1 summarizes the dimensions for the GTI MB gasifier.

Table 6-1
GTI MB Gasifier Overall Dimensions

No. of Trains	2
No. of Gasifiers per Train	1
Gasifier Diameter (ID), ft (top)	8.5
Gasifier Diameter (ID),ft (bottom)	6
GTI Gasifier Height, ft (top)	13.5
GTI Gasifier Height, ft (bottom)	13.5
Refractory Thickness, inches, (top)	6
Refractory Thickness, inches, (bottom)	6
Steam Tube OD, inches	1.9
Gasifier Overall Dimensions:	
Shell Diameter (OD), ft (top)	10
Shell Diameter (OD), ft (bottom)	7.5
Swage Height, ft	3
Total Height, ft	30

6.2.3. Steam Reformer

A conceptual design of the syngas heated steam reformer is shown in Figure 6-3. GTI will provide additional details into the technical development and operational aspects of the HMB steam reformer in another report.

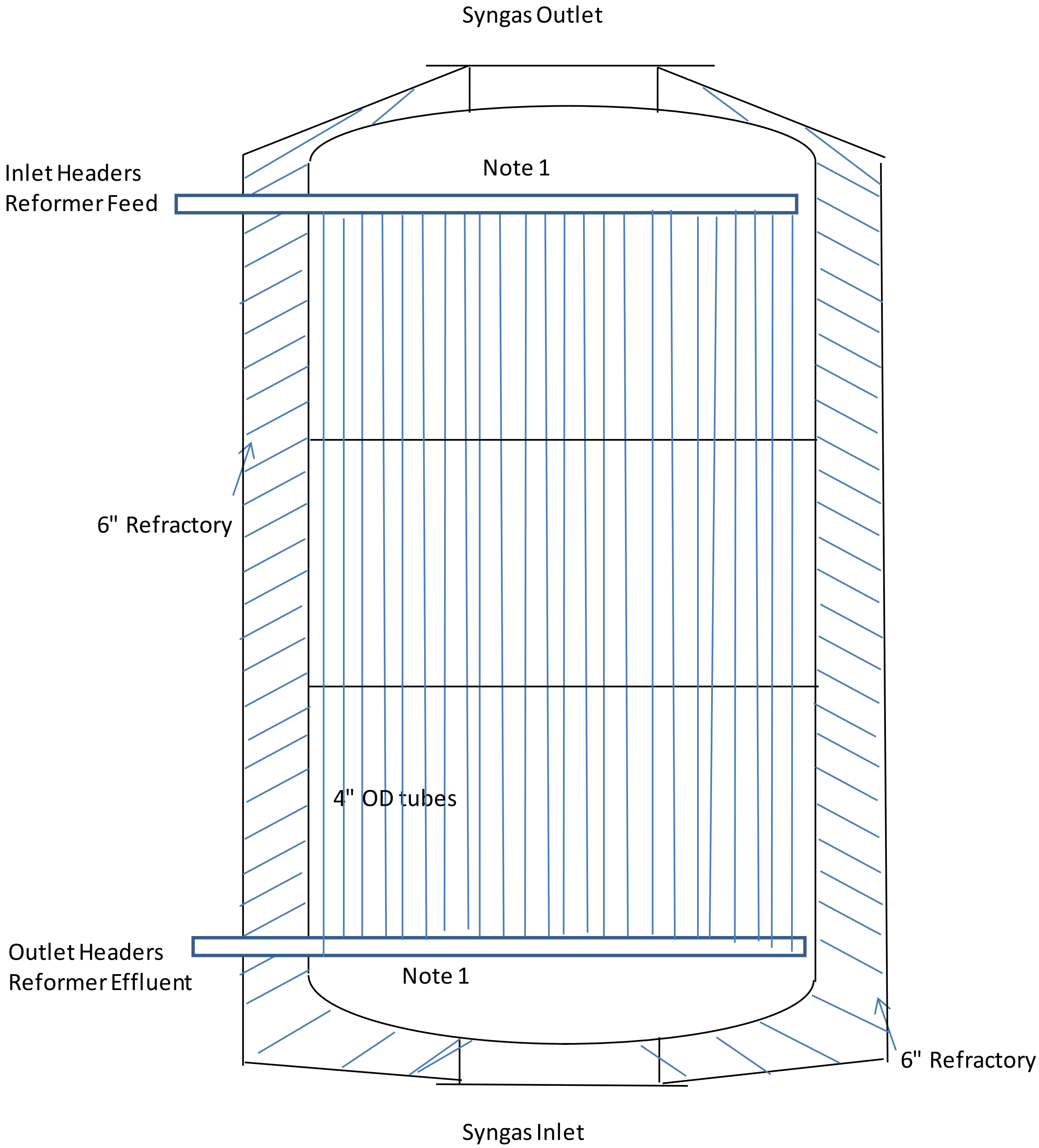
The overall dimension of this reformer is shown in Table 6-2.

The key features of the steam reformer are:

- Shell Side – refractory lined
    - Syngas on shell side; Inlet temperature = 2,600°F, Outlet temperature = 1,800°F
- Tube Side – catalyst filled
    - Steam and natural gas feed; Inlet temperature = 1,085°F, Outlet temperature = 1,500°F
- Reformer duty – 185 MMBtu/h per train, Total reformer duty = 370 MMBtu/h for 2 IGCC trains

The cost of the steam reformer for Case 3 is based on the reformer concept as shown in the conceptual layout in Figure 6-3. The reformer dimensions are estimated from the tube heat exchange surface and catalyst volume requirements based on reforming duty. Incoloy is assumed for the high temperature reforming tube material of construction to provide additional contingency for the reformer cost. Traditional steam reformer tube materials of construction are HK40 or IN-519, which are high Cr and Ni alloys for high temperature service. The reformer vessel wall is assumed to be 316SS construction with 6" of refractory.

Figure 6-3
Syngas Heated Steam Reformer Conceptual Layout



Note 1) Tubes are connected to the headers through pigtails to allow for tube thermal exxpansion

6.2.4. Syngas Cooling & Particulate Filters

The Case 3 syngas cooling / heat integration is optimized to provide reforming duty for the steam reformer, preheating duties for natural gas and reformer steam feeds and high pressure steam generation/superheat. The primary goal is to provide duty for the steam reformer and feed preheat requirements. The remaining cooling duty is used for high pressure steam generation and superheat. Hot

2,600°F syngas generated by the HMB gasifier is heat exchanged with the steam reformer. The cooled syngas exits the reformer at ~1,800°F, hot enough to provide preheating duties for the natural gas feed (900°F) and reformer steam (1,200°F). After feed preheat, the raw syngas is cooled to 685°F by high pressure steam generation before entering the ceramic particulate filters and cyclones. Any remaining particulate matters in the syngas will be removed by these particulate filters and cyclones.

The filtered syngas is then cooled to 450°F through a series of steam generators generating steam for the steam cycle. The cooled syngas enters a water scrubber to remove any chlorides and remaining particulates. The scrubbed syngas is sent to WGS. Part of the scrubber bottoms is used for slag water bath makeup.

6.3. Performance

Table 6-3 shows the power production and auxiliary load breakdown of the Case 3 GTI HMB gasification-based IGCC running a feed mixture of 55% coal/45% natural gas.

Table 6-2
Case 3 Power Generation and Auxiliary Load Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	**GTI Case 3, PRBCC 55%Coal/45% NG 1500F Reformer**
Gas Turbine Power	430,022
Steam Turbine Power	207,376
TOTAL POWER, kWe	**637,397**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	262
Coal Milling	1,402
Slag Handling	250
WTA Coal Dryer Compressor	4,813
WTA Coal Dryer Auxiliaries	318
Natural Gas Compressors	5,078
Gasifier Steam Generator Circ. Pumps	195
Air Separation Unit Auxiliaries	777
Air Separation Unit Main Air Compressor	49,348
Oxygen Compressor	7,547
Nitrogen Compressors	28,026
CO_2 Compressor	23,493
Boiler Feedwater Pumps	4,115
Condensate Pump	219
Quench Water Pump	0
Syngas Recycle Compressor	0
Circulating Water Pump	2,771
Ground Water Pumps	299
Cooling Tower Fans	1,807
Air Cooled Condenser Fans	2,495
Scrubber Pumps	15
Acid Gas Removal	13,740
Gas Turbine Auxiliaries	998
Steam Turbine Auxiliaries	89
Claus Plant/TGTU Auxiliaries	128
Claus Plant TG Recycle Compressor	1,964
Miscellaneous Balance of Plant	1,541
Transformer Losses	2,450
TOTAL AUXILIARIES, kWe	**154,142**
NET POWER, kWe	**483,255**
Net Plant Efficiency, % (HHV)	35.2%
Net Plant Heat Rate, Btu/kWh	9,699
CONDENSER COOLING DUTY, MMBtu/hr	**1,053**
CONSUMABLES	
As-Received Coal Feed, lb/hr	300,991
Thermal Input (Coal+NG), kWt	1,373,649
Raw Water Withdrawal, gpm	3,772
Raw Water Consumption, gpm	3,286

6.4. Elemental Balance

Tables 6-4 and 6-5 show, respectively, the carbon and sulfur balances for the Case 2 co-fired GTI HMB gasifier-based IGCC.

Table 6-3
Case 3 Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
C in Coal Feed	150,701	
C in Natural Gas Feed	67,453	
C in ASU Air	152	
C in Air to Gas Turbine	797	
C in ASU Vent		152
C in Sour Water		-
C in Slag		-
C in Flyash		-
C in Sulfur Product		-
C in Stack Gas		23,322
C in CO2 Product		195,631
Convergence Tolerance		(2)
Total	**219,103**	**219,103**

Table 6-4
Case 3 Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
S in Coal Feed	2,190	
S in Natural Gas Feed	-	
S in ASU Air	-	
S in Air to Gas Turbine	-	
S in ASU Vent		-
S in Sour Water		-
S in Slag		-
S in Flyash		-
Sulfur Product		2,181
Stack Gas		8
S in CO2 Product		-
Convergence Tolerance		0
Total	**2,190**	**2,190**

6.5. Water balance

Water makeup and consumptions are included in the overall utility summary in section 6.8.

6.6. Equipment

The major equipment lists for Case 3 is shown in Table 6-6.

Table 6-5
Case 3 Major Equipment List

VESSELS & TANKS:

Plt No.	Item No.	Item Name	Type	Design Conditions PSIG	Design Conditions deg F	Material of Construction	Quantity per Lot	Units	Inside Diameter Ft	Ht or Tan/Tan Length Ft	Width Ft	Length Ft	Number of Lots	Total Equip Cost $1000
	400C-100	SG Scrubber	Vert	650	450	304Clad shell	2	Vessel	11.0	32.0			2	
	500C-100	Cooled SG KO Drum	Vert	550	450	304Clad shell	2	Vessel	10.0	8.5			2	
	600C-100	Cooled Hydrogenated Tail Gas KO	Vert	15	450	304Clad shell	2	Vessel	4.0	8.0			2	
	600C-101	Hydrogenated Tail Gas Compresso	Vert	15	450	304Clad shell	2	Vessel	3.0	7.5			2	

SHELL & TUBE EXCHANGERS AND AIR COOLERS:

Plt No.	Item No.	Item Name	Type	Design PSIG Shell	Design PSIG Tube	Des Temp, deg F Shell	Des Temp, deg F Tube	Material of Construction Shell	Material of Construction Tube	Duty MMBtu/Hr	Total Bare Tube Area, Ft2	Physical Arrangement In Series	Physical Arrangement In Parallel	Physical Arrangement Total # Req	Total Equip Cost $1000
	300E-107	Process Stm Gen/Quenched SG	Kettle	625	760	1152	580	316SS	316SS	80.7	2,373	1	2	2	
	200E-100	NG Compr Instg Cooler	S&T	281	100	375	375	CS	CS	4.0	1,555	1	2	2	
	200E-101	O2 Compr Instg Cooler	S&T	322	100	375	375	CS	CS	6.2	1,921	1	2	2	
	300E-100	HP Steam Superheater/HT Quenche	S&T	640	2027	1358	1075	316SS	316SS	18.7	1,101	1	2	2	
	300E-102	Process Steam Superheater/HT Qu	Kettle	625	2027	1275	1275	316SS	316SS	49.1	1,445	1	2	2	
	300E-103	NG Preheater 1	S&T	630	762	1215	725	316SS	316SS	13.2	778	1	2	2	
	300E-104	NG Preheater 2	S&T	635	757	1260	975	316SS	316SS	9.7	568	1	2	2	
	400E-103	HP Stm Gen/LTS Feed	Kettle	625	2027	841	706	316SS	316SS	61.4	2,643	1	2	2	
	400E-105	IP Stm Gen/LTS Feed	Kettle	625	2027	635	550	316SS	316SS	27.1	2,196	1	2	2	
	400E-101	SG Scrubber Fd/Btm Exch	S&T	590	635	426	375	CS	CS	2.4	245	1	2	2	
	400E-102	HT WGS Feed/LT WGS SG HX	S&T	735	595	535	504	CS	CS	17.1	4,297	1	2	2	
	400E-202	LP Stm Gen/LTS Elluent	Kettle	100	2027	541	378	CS	CS	20.3	1,220	1	4	4	
	900E-100	HP Stm/Quenched SG Cooler #1	Kettle	625	2027	760	706	316SS	316SS	10.2	651	1	2	2	
	900E-101	IP Stm/Quenched SG Cooler #1	Kettle	625	2027	705	551	316SS	316SS	32.9	1,809	1	2	2	
	900E-102	LP Stm/Quenched SG Cooler #1	Kettle	640	2027	576	376	316SS	316SS	3.1	169	1	2	2	
	300E-101	HP Steam Generator/HT Quenched	Kettle	625	2027	705	705	316SS	316SS	66.0	1,940	1	2	2	
	400E-203	CW/LTS Effluent	S&T	543	100	395	375	316SS	CS	67.8	3,985	1	2	2	
	400E-204	Condensate Preheat/LTS Effluent	S&T	635	100	398	375	CS	CS	125.8	12,009	1	2	2	
	500E-100	CO2 Compressor 1st Stg Cooler	S&T	275	100	375	375	304SS	CS	2.4	829	1	2	2	
	500E-101	CO2 Compressor 2nd Stg Cooler	S&T	550	100	375	375	304SS	CS	2.4	554	1	2	2	
	500E-102	CO2 Compressor 3rd Stg Cooler	S&T	1056	100	375	375	304SS	CS	2.4	563	1	2	2	
	500E-104	SC CO2 Cooler	S&T	549	100	851	375	304SS	CS	2.4	142	1	2	2	
	600E-100	Sulfur Cooler	S&T	18	100	425	155	316SS	CS	0.1	9	1	2	2	
	600E-101	Hydorgenater Tail Gas Cooler	S&T	11	100	625	155	CS	CS	3.1	183	1	2	2	
	600E-102	Hydrogenated Tail Gas Compresso	S&T	70	100	594	155	CS	CS	1.9	114	1	2	2	
	600E-103	Hydrogenated Tail Gas Compresso	S&T	528	100	620	155	CS	CS	1.5	96	1	2	2	
	700E-100	GT Feed Superheater	S&T	735	500	580	460	CS	CS	38.5	2,265	1	2	2	
	100E-100	ASU Air Compr 1st Instg Cooler	S&T	322	100	375	375	CS	CS	6.6	1,123	1	2	2	
	100E-101	ASU Air Compr 2nd Instg Cooler	S&T	322	100	375	375	CS	CS	7.5	1,170	1	2	2	
	100E-102	ASU N2 Primary Compr Instg Cool	S&T	322	100	376	375	CS	CS	5.2	749	1	2	2	

NOTES:

REV	DATE	REVISIONS	PROC. ENG	UNIT ENG
0		Issued for Phase 1 Report	AKL	

DOE/NETL Advanced Gasification Technologies Program
GTI Hybrid Molten Bed Gasifier for Production of High Hydrogen Syngas Project (GTI HMB Gasifier)
MAJOR EQUIPMENT LIST

JOB NUMBER: A02220
DRAWING No.: DS-EQUIP-001
REV.: 0
Page 1 of 2

COMPRESSORS, BLOWERS & DRIVERS:
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Plt No.	Item No.	Item Name	Type	Design Conditic PSIG	deg F	Material of Construction Wheel or Impel'r	Casing	Design Capacity Des Flow SCFM	Inlet PSIA	Delta P PSI	Comp BHP	Driver HP	Type	Total # Req	Total Equip Cost $1000
	200K-100A	NG Compressor 1st Stage	Cent	365	328	CS	CS	17,044	99.5	250.5	1,983	2181	Motor	2	
	200K-100B	NG Compressor 2nd Stage	Cent	741.7	240	CS	CS	17,044	344.5	382.2	1,111	1222	Motor	2	
	200K-101A	O2 Compressor 1st Stage	Cent	322.144	329	CS	CS	27,288	124.5	182.6	2,310	2541	Motor	2	
	200K-101B	O2 Compressor 2nd Stage	Cent	764.7	317	CS	CS	27,288	301.644	448.1	2,289	2518	Motor	2	
	600K-100	Hydrogenator TG Compressor 1st	Cent	505	992	CS	CS	2,173	10.1	479.9	1,197	1316	Motor	2	
	600K-101	Hydrogenator TG Compressor 2nd	Cent	505	523	CS	CS	1,957	65	425.0	465	511	Motor	2	
	100K-100A	1st Stg ASU Air Compressor	Cent	52	302	CS	CS	66,871	12.9	24.1	6,180	6798	Motor	4	
	100K-100B	2nd Stg ASU Air Compressor	Cent	95	326	CS	CS	60,033	31.5	48.5	5,184	5702	Motor	4	
	100K-100C	3rd Stg ASU Air Compressor	Cent	205	324	CS	CS	60,033	74.5	115.5	5,202	5722	Motor	4	
	500K-100A	1st Stg CO2 Compressor	Cent	265	189	CS	CS	26,019	148.7	101.3	1,122	1235	Motor	4	
	500K-100B	2nd Stg CO2 Compressor	Cent	515	242	CS	CS	35,289	234.5	265.5	2,344	2578	Motor	4	
	500K-100C	3rd Stg CO2 Compressor	Cent	1020	239	CS	CS	35,289	494.5	510.5	2,191	2410	Motor	4	
	100K-101A	1st Stg Primary N2 Compressor	Cent	163	351	CS	CS	42,588	55.9	92.1	3,939	4333	Motor	4	
	100K-101B	2nd Stg Primary N2 Compressor	Cent	400	341	CS	CS	42,588	142.5	242.5	3,958	4354	Motor	4	
	100K-102	Secondary N2 Compressor	Cent	400	284	CS	CS	12,428	181.5	203.5	843	928	Motor	2	

PUMPS & DRIVERS:
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Plt No.	Item No.	Item Name	Type	Design Conditi PSIG	deg F	Material of Construction Wheel or Impel'r	Casing	Design Capacity Des Flow GPM	Inlet PSIG	Delta P PSI	Pump BHP	Driver HP	Type	Total # Req	Total Equip Cost $1000
	300G-100	SG Steam Gen BFW Pump	Cent.	790	450	304SS	304SS	292	590	149	32	35	Motor	4	
	400G-100	SG Scrubber Recirc Pump	Cent.	630	450	304SS	304SS	308	560	21	5	5	Motor	4	
	500G-100	SC CO2 Pump	Cent.	2317	450	304SS	304SS	988	995	1212	873	970	Motor	4	

PACKAGED & MISC EQUIPMENT:
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Plt No.	Item No.	Item Name	Type	Design Conditions PSIG	deg F	Mat Of Construct	Design Capacity	Remarks	Total # Req	Total Equip Cost $1000
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		Particulate Filters				304SS		Capacity Factored		

6.7. Utilities

Table 6-7 shows the utilities summary of the Case 3.

Table 6-6
Case 3 Utilities Summary

Item Name	Elect. Power	Steam							Water Requirement, 1000 Lbs/Hr													Cooling Water	
		1000 Lbs/Hr							Cold Cond	Return Cond	BFW	LP Cond	MP Cond 1	MP Cond 2	Process Effluent	SWS Stripped Water	Open	Open	Raw Makeup	Blowdown	Wastewater to Treatment		
	KW	SH Process Stm: 715 PSIA/ 1200F	SHHP: 1815 PSIA/ 1000F	HP: 1875 PSIA/ 627F Sat	Process Stm: 720 PSIA / 505 F Sat	IP: 525 PSIA / 472 F Sat	250 PSIA / 786F	LP: 65 PSIA / 298 F Sat	Cold Cond / 90F	Return Cond / 235F	1975 PSIA / 288F	LP Cond/ 293F	MP Cond/ 471F	250 PSIA / 401F					Raw Makeup Water	Blowdown	Waste W / @ 100 F	CW, MMbtu/hr	C.W. circ. GPM
PROCESS/GASIFICATION ISLAND																							
COAL/SLAG HANDLING & MILLING:																							
COAL HANDLING	262	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
COAL MILLING	1,402	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SLAG HANDLING	250	-	-	-	-	-	-	-	-	-	-	-	-	-	-	28	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
WTA DRYING:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
WTA COAL DRYER COMPRESSOR	4,813	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
WTA COAL DRYER AUXILIARIES	318	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	8	821
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
AIR SEPARATION UNIT:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ASU AUXILIARIES	777	-	-	-	-	-	23	38	-	-	-	(38)	(23)	-	-	-	-	-	-	-	-	-	-
ASU MAIN COMPRESSOR	49,348	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ASU INTERCOOLER (incl O2 & N2 Intercooling)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	160	15,980
OXYGEN COMPRESSOR	7,547	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	12	1,247
NITROGEN COMPRESSOR	28,027	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NITROGEN BOOST COMPRESSOR	456	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM REFORMER:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NATURAL GAS COMPRESSOR	5,078	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NG COMPRESSOR INTERCOOLERS	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	8	808
REFORMING STEAM INJECTION	-	254	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GASIFIER & SYNGAS COOLING:	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GTI GASIFIER WALL STEAM GENERATORS	-	-	-	-	(93)	-	-	-	-	-	94	-	-	-	-	-	-	-	-	(0)	-	-	-
SG / PROCESS STEAM GENERATOR	-	-	-	-	(160)	-	-	-	-	-	161	-	-	-	-	-	-	-	-	(1)	-	-	-
SG / PROCESS STEAM SUPERHEATER	-	(254)	-	-	254	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SG / HP STEAM GENERATOR	-	-	-	(137)	-	-	-	-	-	-	137	-	-	-	-	-	-	-	-	(1)	-	-	-
SG / HP STEAM SPERHEATER	-	-	(137)	137	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GASIFIER STM GEN CIRC PUMP	195	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ASH COOLING	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	15	1,481
QUENCH COOLER	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
QUENCH WATER PUMP	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS RECYCLE COMPRESSOR	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
QUENCHED SYNGAS COOLER/STEAM GEN	-	-	-	(23)	-	(69)	-	(8)	-	-	100	-	-	-	-	-	-	-	-	(1)	-	-	-
IP STEAM LETDOWN	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SCRUBBER, SHIFT & SYNGAS COOLING:	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS SCRUBBER	-	-	-	-	-	-	-	-	62	-	-	-	-	-	(63)	-	-	-	-	-	-	-	-
SCRUBBER PUMPS	15	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SHIFT REACTOR	-	-	-	249	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
HT SHIFT SYNGAS COOLER/STEAM GEN	-	-	-	(138)	-	(57)	-	-	-	-	196	-	-	-	-	-	-	-	-	(1)	-	-	-
LT SHIFT SYNGAS COOLER/STEAM GEN	-	-	-	-	-	-	-	(87)	1,415	(1,415)	88	-	-	-	-	-	-	-	-	0	-	-	-
LT SHIFT SYNGAS CW COOLER	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	169	16,918
LT SHIFT SYNGAS KO DRUM	-	-	-	-	-	-	-	-	-	-	-	-	-	-	(239)	-	-	-	-	-	-	-	-
SOUR WATER STRIPPER	-	-	-	-	-	-	-	16	-	-	-	(16)	-	-	308	(271)	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ACID GAS REMOVAL:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
DOUBLE STAGE SELEXOL	13,740	-	-	-	-	-	-	108	-	-	-	(108)	-	-	-	-	-	-	-	-	-	-	-
DOUBLE STAGE SELEXOL COOLING	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	46	4,548
CLAUS PLANT/TGTU AUXILIARIES	128	-	-	-	-	(3)	-	(2)	-	-	8	-	(3)	-	-	-	-	-	-	-	-	-	-
CLAUS PLANT TG RECYCLE COMPRESSOR	1,224	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	13	1,305
CLAUS PLANT TG RECYCLE COMPRESSOR INSTG KO DRUMS	-	-	-	-	-	-	-	-	-	-	-	-	-	-	(4)	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2 COMPRESSION:	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2 COMPRESSORS	23,493	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO2 COMPRESSION INTERCOOLER	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	108	10,805
CO2 COMPRESSION KO DRUMS	-	-	-	-	-	-	-	-	-	-	-	-	-	-	(1)	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS REHEAT:	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SYNGAS REHEATER	-	-	-	-	-	36	-	55	-	-	-	(55)	(36)	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MISCELLANEOUS	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
BALANCE OF PLANT	3,000	-	-	-	-	-	-	4	-	-	-	(4)	-	-	-	-	-	-	-	-	-	-	-
	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SUBTOTAL PROCESS/GASIFICATION ISLAND	**140,074**	**-**	**(137)**	**89**	**-**	**(93)**	**23**	**124**	**1,478**	**(1,415)**	**784**	**(221)**	**(62)**	**-**	**(0)**	**(243)**	**-**	**-**	**-**	**(3)**	**-**	**540**	**53,913**
POWER ISLAND & STEAM CYCLE																							
GAS TURBINE:																							
GAS TURBINE GENERATOR	(430,022)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
GAS TURBINE AUXILIARIES	(998)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM TURBINE:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM TURBINE GENERATOR	(207,376)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM TURBINE AUXILIARIES	(89)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MISCELLANEOUS:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TRANSFORMER LOSSES	2,450	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM CYCLE:		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STEAM / CONDENSATE IMPORT		-	137	(89)	-	130	-	97	(1,478)	1,415	-	221	62	-	-	-	-	-	572	3	-	-	-
STEAM / CONDENSATE (EXPORT)		-	-	-	-	(36)	(23)	(221)	-	-	(784)	-	-	-	-	-	-	-	-	-	6	-	-
AIR COOLED CONDENSER FANS	2,495	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CW COOLED SURFACE CONDENSER		-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	527	52,619
CONDENSATE PUMPS	219	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
BOILER FEED WATER PUMPS	4,115	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SUBTOTAL POWER ISLAND & STEAM CYCLE	**(629,206)**	**-**	**137**	**(89)**	**-**	**93**	**(23)**	**(124)**	**(1,478)**	**1,415**	**(784)**	**221**	**62**	**-**	**-**	**-**	**-**	**-**	**572**	**3**	**6**	**527**	**52,619**
COOLING WATER & COOLING TOWER																							
COOLING TOWER:																							
COOLING WATER PRODUCTION		-														335.0			1,260.2		(361.6)	(1,294)	(129,277)
COOLING TOWER FANS	2,576	-																					
GROUND WATER PUMPS	299	-																					
CIRCULATING WATER PUMP	1,679	-																					
SUBTOTAL CW & CT	**4,554**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**-**	**335**	**-**	**-**	**1,260**	**-**	**(362)**	**(1,294)**	**(129,277)**
GRAND TOTAL IGCC	**(484,578)**	**-**	**0**	**0**	**-**	**0**	**0**	**(0)**	**-**	**-**	**0**	**0**	**(0)**	**-**	**(0)**	**92**	**-**	**-**	**1,832**	**(0)**	**(356)**	**(228)**	**(22,744)**

6.8. Capital Cost

6.8.1. Total Plant Cost

Table 6-8 shows the total plant cost (TPC) summary of GTI MB Case 3 compared to the Reference Bench Mark case.

Table 6-7
Total Plant Cost Summary

Total Plant Cost (June 2011)		GTI HMB Case 3	Shell Reference Case Bench Mark (Case 2a)
Acct. No.	Item/Description	$MM	$MM
1	COAL & SORBENT HANDLING	32.7	49.4
2	COAL & SORBENT PREP & FEED	153.2	237.8
3	FEEDWATER & MISC BOP SYSTEMS	29.3	34.4
4	GASIFIER & ACCESSORIES	453.4	751.4
5	GAS CLEANUP & PIPING	268.8	289.9
5B.2	CO2 Compression & Drying	51.1	66.3
6	COMBUSTION TURBINE/ACCESSORIES	159.4	159.4
7	HRSG, DUCTING & STACK	54.3	54.0
8	STEAM TURBINE GENERATOR	114.1	122.5
9	COOLING WATER SYSTEM	25.2	27.0
10	ASH/SPENT SORBENT HANDLING SYS	26.3	44.4
11	ACCESSORY ELECTRIC PLANT	97.6	105.0
12	INSTRUMENTATION & CONTROL	31.1	32.0
13	IMPROVEMENTS TO SITE	22.0	22.5
14	BUILDINGS & STRUCTURES	20.5	20.9
	CALCULATED TOTAL COST	1,539.0	2,016.9

6.9. Operating Costs

Table 6-6 summarizes the operating costs for GTI MB Case 3 compared to the Reference Bench Mark case.

Table 6-8
Operating Cost Breakdown

OPERATING COSTS, 2011 $MM/yr	GTI MB Case 3	Shell Reference Case Bench Mark (Case 2a)
FIXED OPERATING COSTS		
Annual Operting Labor Cost	$7.2	$7.2
Maintainence Labor Cost	$14.9	$19.5
Administration & Support Labor	$5.5	$6.7
Property Taxes and Insurance	$30.8	$40.3
TOTAL FIXED OPERATING COSTS	$58.4	$73.7
VRIABLE OPERATING COSTS (@100% CF)		
NON-FUEL VARIABLE OPERATING COSTS		
Maintenance Material Cost	$34.6	$45.3
Water	$1.7	$1.5
Chemicals		
MU & WT Chemicals	$1.6	$1.5
Other Chemicals & Catalysts	$1.7	$2.6
Waste Disposal	$2.7	$5.4
TOTAL NON_FUEL VARIABLE OPERATING COSTS	$42.3	$56.3
FUEL (@100% CF)		
Coal	$25.9	$50.4
Natural Gas	$95.7	$0.0
TOTAL VARIABLE OPERATING COSTS	$163.9	$106.7

6.10. Cost of Electricity

Table 6-7 shows a summary of the power output, CAPEX, OPEX, COE and cost of CO_2 capture for Case 3.

Table 6-9
Operating Cost Breakdown

	GTI MB Case 3	Shell Reference Case Bench Mark (Case 2a)
CAPEX, $MM		
Total Installed Cost (TIC)	$1,155	$1,512
Total Plant Cost (TPC)	$1,539	$2,017
Total Overnight Cost (TOC)	$1,902	$2,472
OPEX, $MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC_{Fix})	$58	$74
Variable Operating Cost Less Fuel (OC_{VAR})	$42	$56
Fuel Cost (OC_{Fuel})	$122	$50
Power Production, Mwe		
Gas Turbine	430.0	430.0
Steam Turbine	207.4	222.2
Auxiliary Power Consumption	154.1	192.6
Net Power Output	483.3	459.6
Power Generated, MWh/yr (MWH)	4,233,318	4,026,096
COE, excl CO2 TS&M, mills/kWh	125.8	144.8
COE, incl CO2 TS&M, mills/kWh	140.6	165.6
Cost of CO2 Avoided excl CO2 TS&M, $/ton CO2	$54.1	$79.2
Cost of CO2 Avoided incl CO2 TS&M, $/ton CO2	$71.3	$104.4

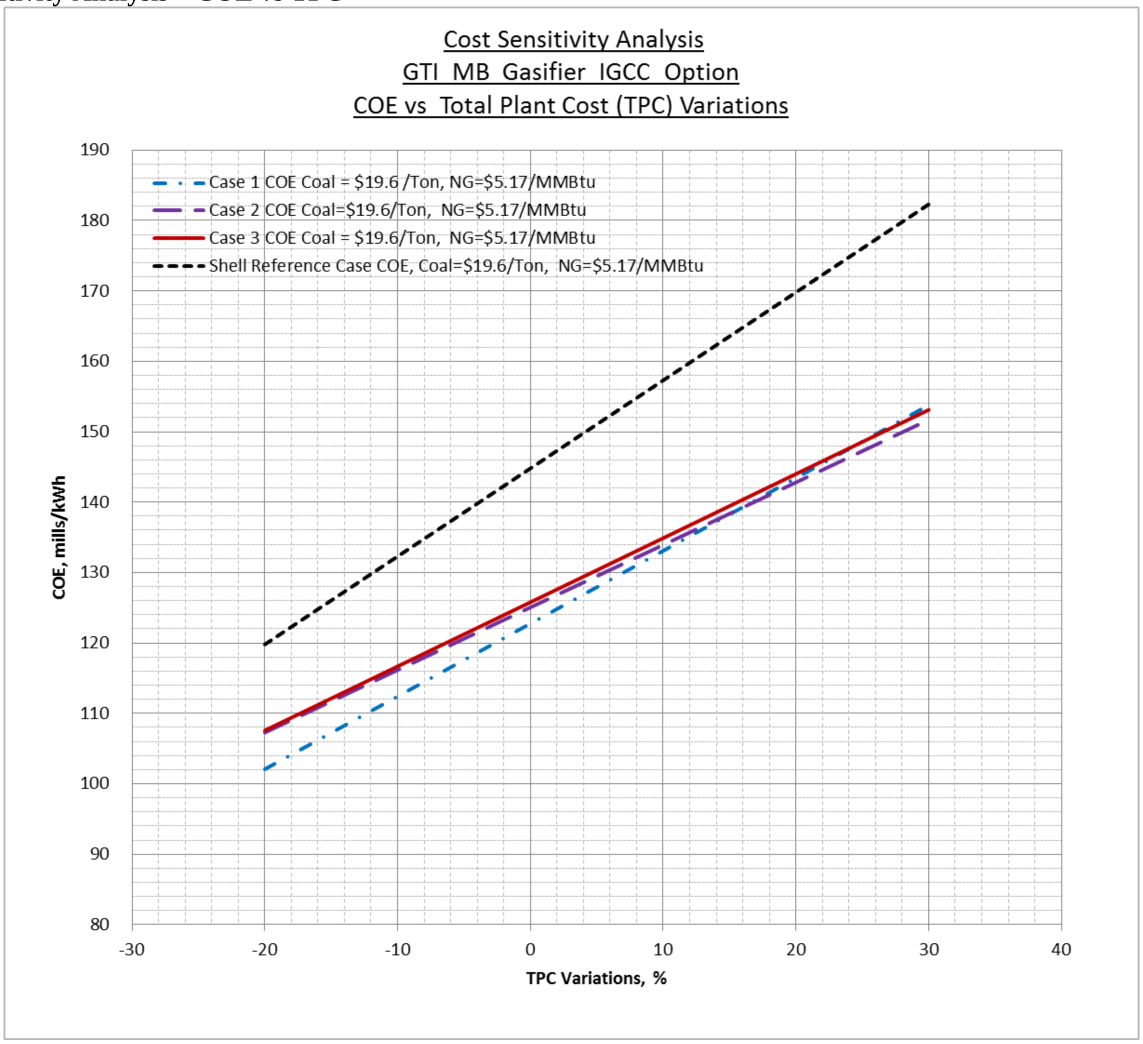
7. SENSITIVITY ANALYSIS

Sensitivity analysis was carried out to determine the effects of various parameters on the overall IGCC COE. The parameters investigated here include: TPC, feedstock prices, capacity factor, CO_2 sales price and cost of CO_2 emissions.

7.1. Total PLANT Cost (TPC)

Figure 7-1 shows IGCC COEs variation with TPC from -20% to +30%.

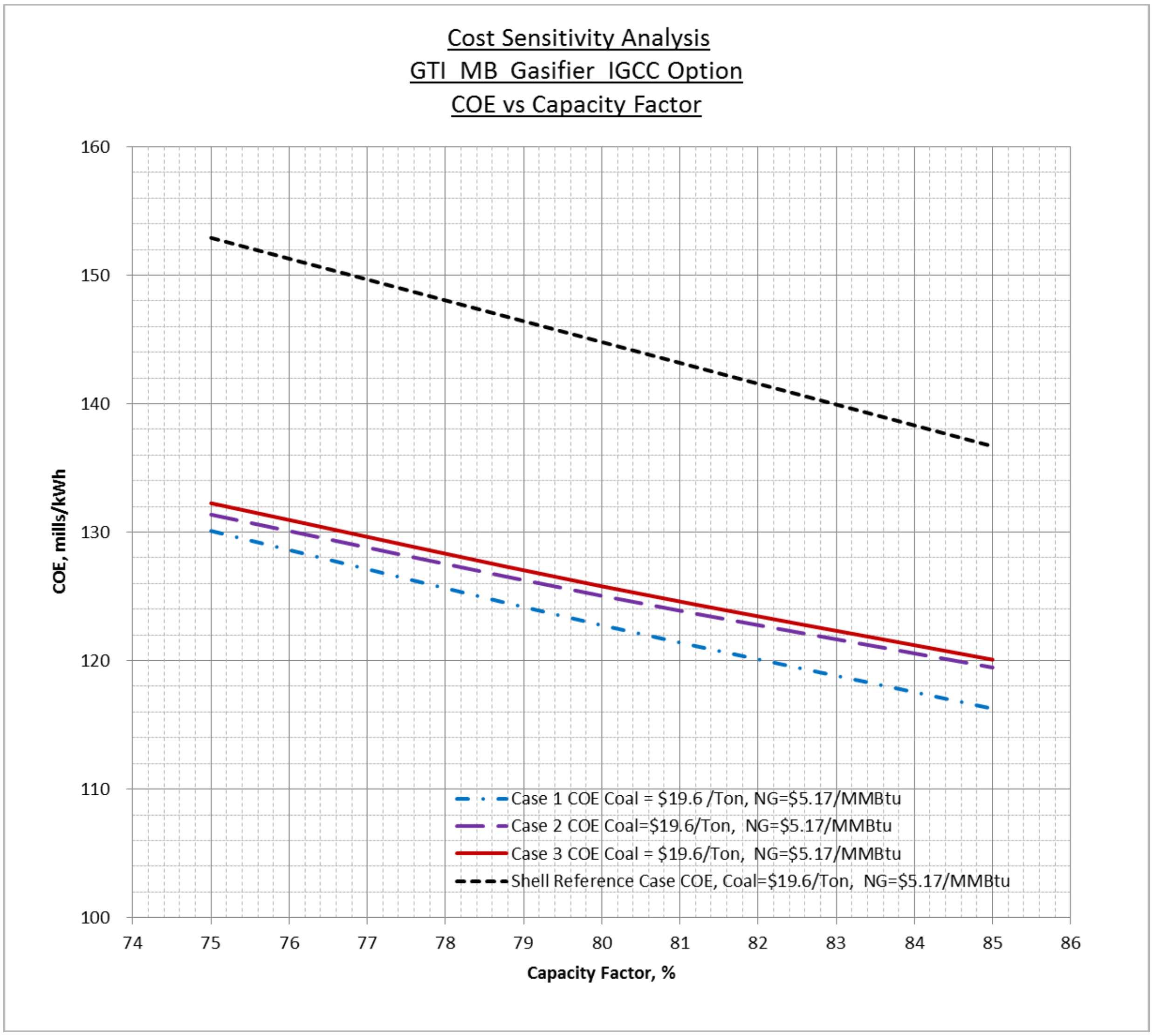
Figure 7-1
Sensitivity Analysis – COE vs TPC



7.2. Capacity Factor

The baseline IGCC plant capacity factor used in this study is 80%. Figure 7-2 shows how the IGCC COE varies with plant capacity factor as it varies from 75% to 85%.

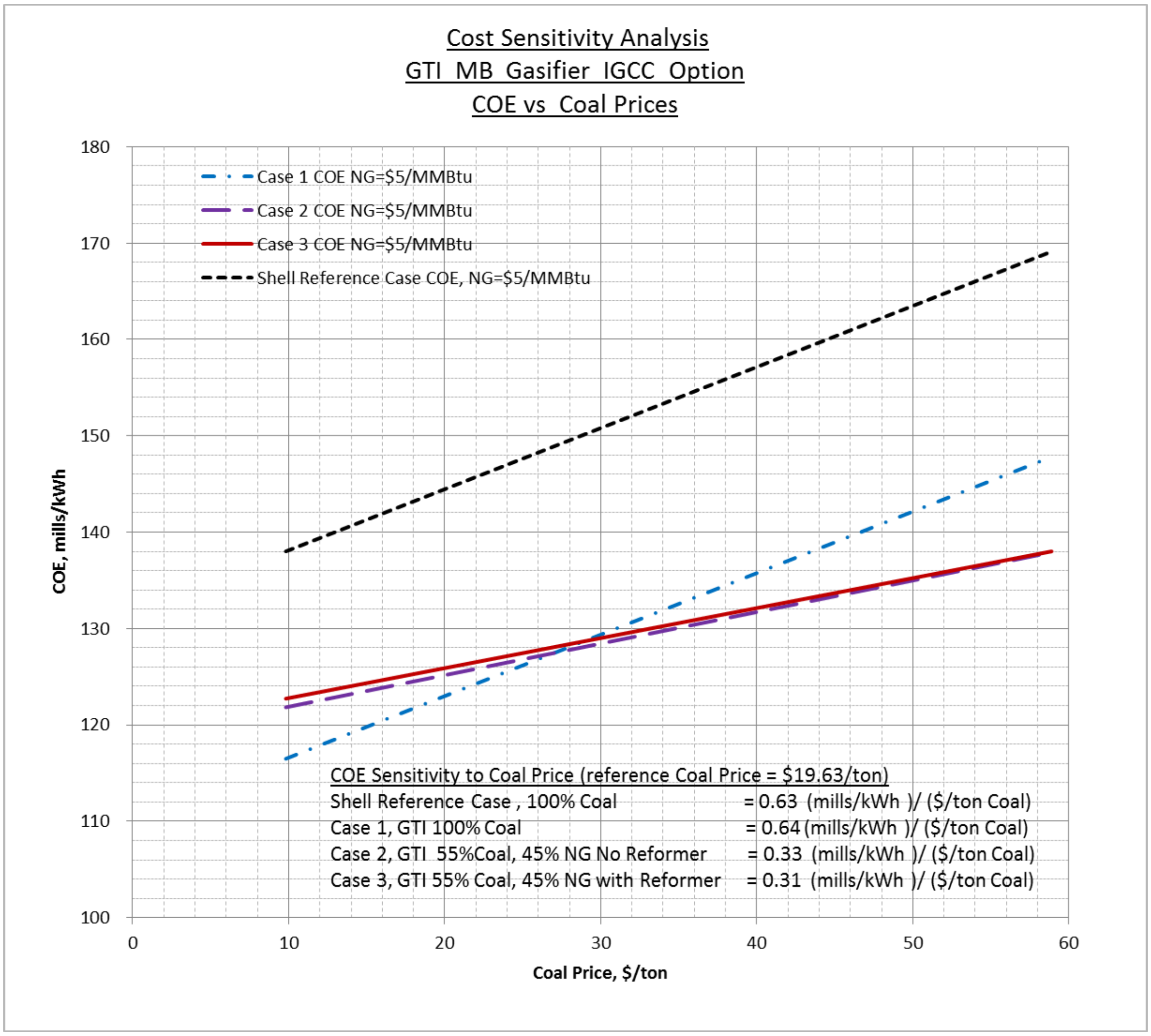
Figure 7-2
Sensitivity Analysis – COE vs IGCC Plant Capacity Factor



7.3. Coal Price

The baseline IGCC plant coal price used in this study is $19.6/Ton. Figure 7-3 shows how the IGCC COE varies with coal price as it varies from -25% to +300% (~$10/Ton to ~$60/Ton).

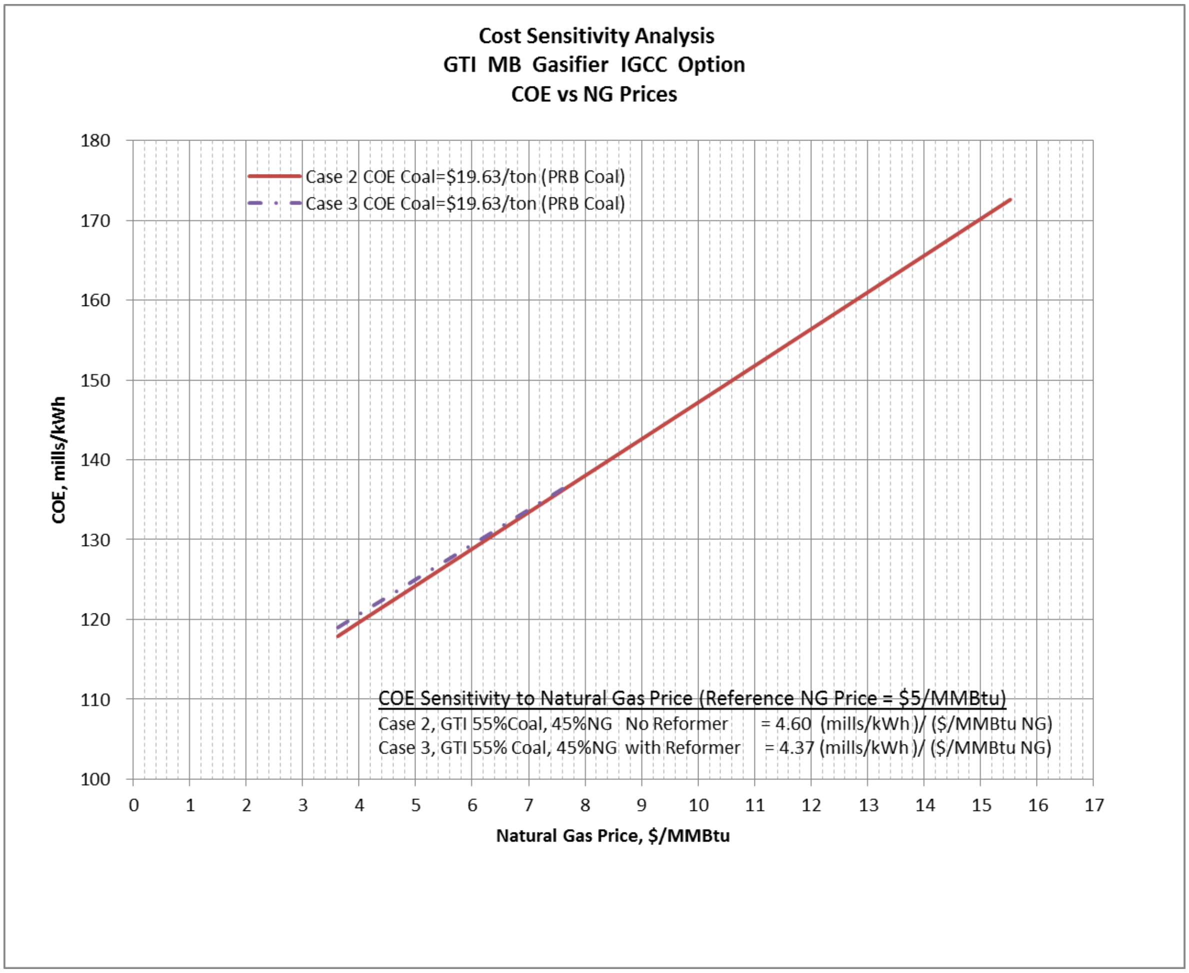
Figure 7-3
Sensitivity Analysis – COE vs Coal Price



7.4. Natural Gas Price

The baseline IGCC plant natural gas price used in this study is \$5.34/ 1000 ft3 (\$5.17/MMBtu). Figure 7-4 shows how the IGCC COE varies with natural gas price as it varies from -25% to +300% (~\$4/MMBtu to ~\$15/MMBtu).

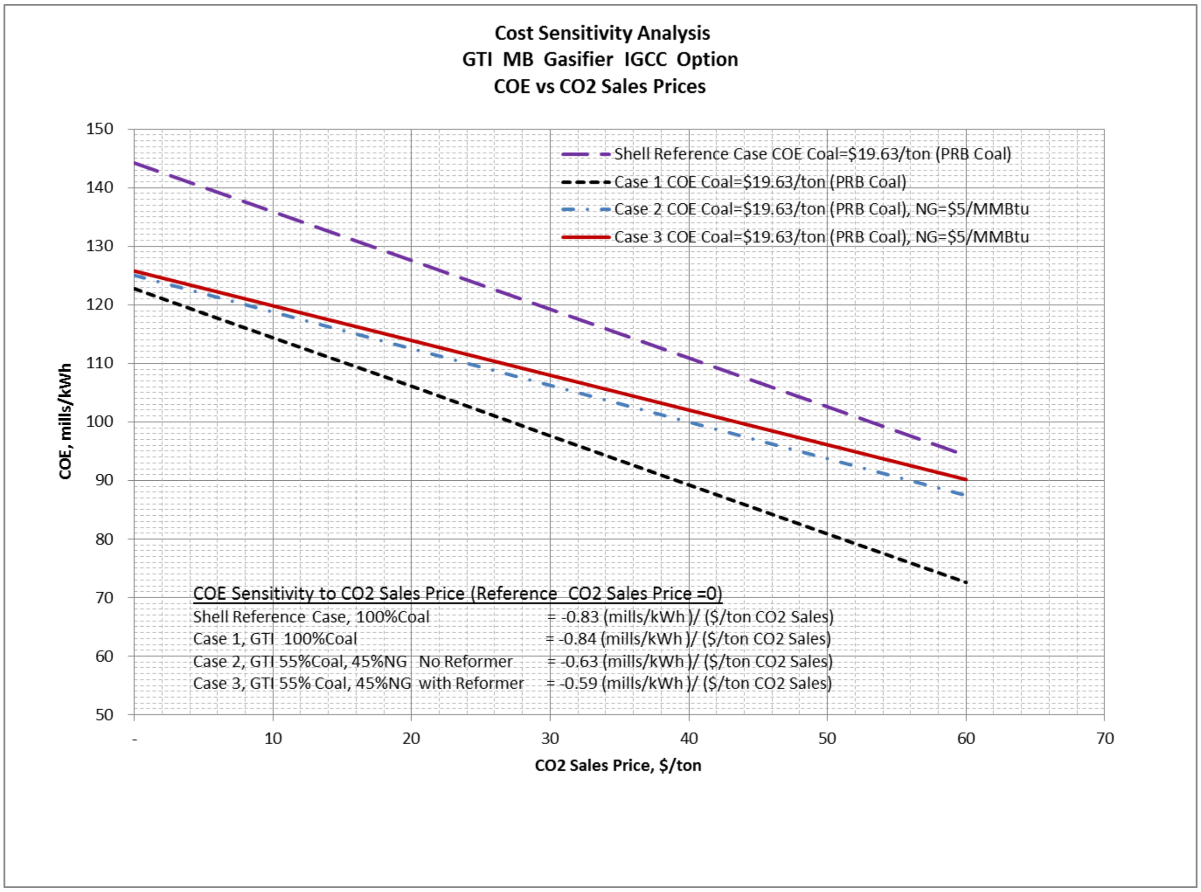
Figure 7-4
Sensitivity Analysis – COE vs Natural Gas Price



7.5. CO_2 Sales Price

Sensitivity to CO_2 sales at plant gate prices is shown in Figure 7-5. The baseline case assumes that the CO_2 product carries no value ($0/ton). The sales price is subsequently varied to a maximum of $60/ton to determine its effect on the IGCC plant's COE.

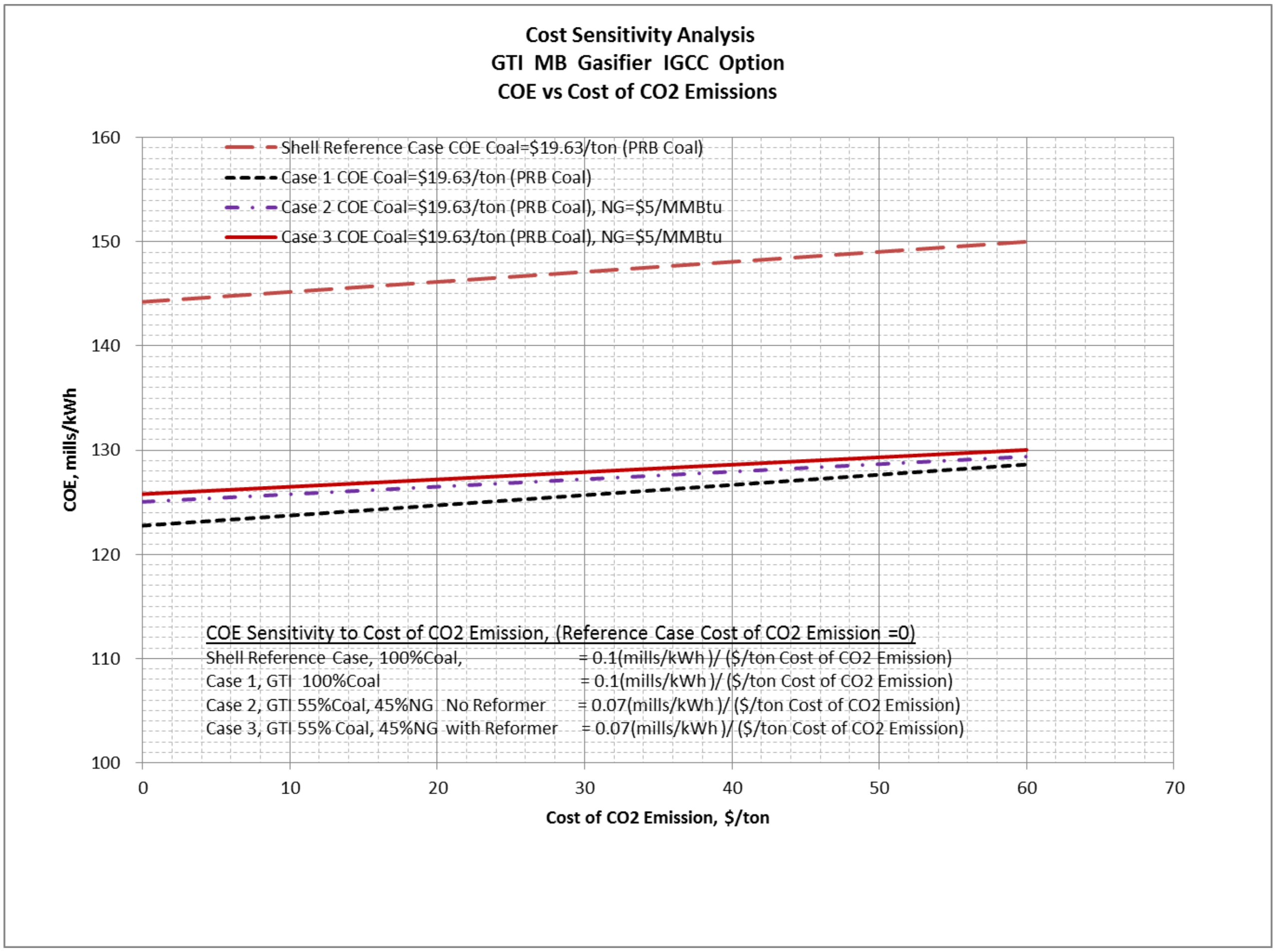
Figure 7-5
Sensitivity Analysis – COE vs CO_2 Sales Price



7.6. Cost of CO_2 Emissions

The sensitivity to CO_2 emissions costs is shown in Figure 7-6. The baseline case assumes that there are no costs associated with venting CO_2 to the atmosphere ($0/ton). The cost of CO_2 emissions is subsequently varied to a maximum of $60/ton to determine its effect on the IGCC plant's COE.

Figure 7-6
Sensitivity Analysis – COE vs Cost of CO_2 Emissions



8. CONCLUSIONS AND RECOMMENDATIONS

8.1. Conclusions

The objective of this techno-economic analysis is to assess the cost and performance of an IGCC power plant with CO_2 capture that utilizes GTI's hybrid molten bed (HMB) gasification process to gasify low rank Montana PRB coal. The GTI HMB gasifier is a dual coal-natural gas fueled molten bed gasification process. By varying coal and natural gas feed rates, and steam to natural gas ratio to the gasifier, the syngas H_2/CO ratio can be optimized for producing electricity by IGCC with reduced water gas shift requirements.

Three GTI HMB Gasifier cases with variations in feed mix, gasifier configuration and heat integration schemes are analyzed for the IGCC plant with CO_2 capture option. These cases are evaluated against a reference Case S1B from the DOE/NETL 1399 Baseline Study. The reference Case S1B is a Shell SCGP gasifier based IGCC power plant with CO_2 capture. Schematic depictions of these cases are included in the simplified block flow diagrams in figures 2-1 to 2-4 in section 2.

- Reference Case – Shell SCGP Gasifier, 100% PRB Coal Feed
- Case 1- GTI HMB Gasifier, 100% PRB Coal Feed
- Case 2- GTI HMB Gasifier, 55% PRB Coal / 45% NG Feed
- Case 3- GTI HMB Gasifier, 55% PRB Coal / 45% NG Feed with Steam Reformer

Tables 8-1 and 8-2 summarize the plant performance results for each case. Table 8-3 summarizes the plant economic results for each case.

Table 8-1
Plant Performance Summary

Case	Reference Case	Case 1	Case 2	Case 3
Source	Nexant Modeling	Nexant Modeling	Nexant Modeling	Nexant Modeling
Gasifier & Coal Feed Technology	Shell SCGP Gasifier	GTI HMB Gasifier	GTI MBG	GTI Hybrid MBG
Coal Type	PRB	PRB	PRB	PRB
Feed Mix, % HHV	100% Coal	100% Coal	55% Coal / 45% NG	55% Coal / 45% NG
As-Received Coal Feed, lb/hr	585,971	580,414	334,168	300,991
Natural Gas Feed, lb/hr	-	-	103,680	93,386
Carbon Capture, %	90	90	90	90
Cold Gas Efficiency, %	80.6	81.6	78.0	87.4
Acid Gas Recovery Technology	Selexol	Selexol	Selexol	Selexol

Case	Reference Case	Case 1	Case 2	Case 3
Power Summary, MWe				
Power Generation :				
Gas Turbine	*430.0*	*432.1*	*431.3*	*430.0*
Steam Turbine	*222.2*	*211.1*	*267.6*	*207.4*
Total Gross Power	*652.2*	*643.2*	*698.9*	*637.4*
Auxiliary Load Total	*192.6*	*188.7*	*189.0*	*154.1*
Net Power Generation	459.6	454.5	509.9	483.3
Net Plant Efficiency, % HHV	31.2%	31.2%	33.4%	35.2%

Table 8-2
Plant Economic Summary

Case	Reference Case	Case 1	Case 2	Case 3
Capacity Factor (CF), %	80	80	80	80
Net Power Generation, MWe	459.6	454.5	509.9	483.3
2011 Capital Cost, $MM				
Total Plant Cost, $MM	2,017	1,643	1,584	1546
Total Overnight Cost, $MM	2,472	2,012	1,938	1910
Total Plant Cost/kW, $/kW	4,389	3,615	3,106	3,199
2011 Operating Cost, $MM/yr				
Fixed Operating Costs	*74*	*62*	*60*	*58*
Variable Operating Costs @ 100% CF	*56*	*49*	*44*	*42*
Fuel Costs @ 100% CF, Coal @$19.63/ton	*50*	*50*	*29*	*26*
NG @ $5/MMBtu	*0*	*0*	*106*	*96*
Cost of Electricity (excl TS&M), mills/kWh	144.8	122.8	125.0	125.8
Cost of Electricity (incl TS&M), mills/kWh	165.6	143.7	140.7	140.6

Table 8-3
Total Plant Summary by Account

Code of Accounts	TOTAL PLANT COST, 2011 $MM	Reference Case	Case 1	Case 2	Case 3
1	COAL & SORBENT HANDLING	49.4	49.1	34.9	32.7
2	COAL & SORBENT PREP & FEED	237.8	236.3	164.2	153.2
3	FEEDWATER & MISC BOP SYSTEMS	34.4	29.9	36.1	29.3
4	GASIFIER & ACCESSORIES	751.4	401.5	432.2	453.4
5A	GAS CLEANUP & PIPING	289.9	289.4	269.7	268.8
5B	CO2 REMOVAL & COMPRESSION	66.3	65.6	56.2	51.1
6	COMBUSTION TURBINE/ACCESSORIES	159.4	159.4	159.4	159.4
7	HRSG, DUCTING & STACK	54.0	53.5	54.2	54.3
8	STEAM TURBINE GENERATOR	122.5	112.1	140.6	114.1
9	COOLING WATER SYSTEM	27.0	27.9	28.3	25.2
10	ASH/SPENT SORBENT HANDLING	44.4	39.6	28.1	26.3
11	ACCESSORY ELECTRIC PLANT	105	104.1	105.7	97.6
12	INSTRUMENTATION & CONTROL	32.0	31.9	31.9	31.1
13	IMPROVEMENTS TO SITE	22.5	22.2	22.1	22
14	BUILDINGS & STRUCTURES	20.9	20.9	20.7	20.5
	TOTAL TPC	2,016.9	1643.2	1584.2	1539.0

8.2. Energy Efficiency & Plant Performance

The net plant efficiency (NPE) on a HHV basis for the reference and the GTI HMB cases are shown in Table 8-1. The NPE for Cases 2 and 3 which are based on the coal/NG co-feed GTI HMB gasifiers are 33.4% and 35.2% respectively. The NPE for 100% coal feed reference Shell gasifier IGCC case and the similar GTI HMB Case 1 are both 31.2%.

The NPE for Cases 3 is the highest among the four cases at 35.2%. The configuration of the Case 3 gasification process is made more efficient than the other IGCC cases by:

- recuperating heat from its walls
- recovering heat from the hot, raw syngas through endothermic steam reforming of natural gas, enabling chemical energy to be returned as fuel to the gasifier (syngas is cooled from 2,600°F to 1,806°F)
- heat recycle to the gasifier through natural gas and steam preheating

The cold gas efficiency (CGE) is a measure of the conversion efficiency of feed to H_2 + CO syngas products based on their HHV. In equation form it is as follows:

- $CGE = (H_2+CO)_{HHV} / (Feed)_{HHV} * 100$ (equation 1)

The CGE for the reference and the GTI HMB cases are shown in Table 8-1.

The CGE for 100% coal feed reference Shell gasifier IGCC case and the similar GTI HMB Case 1 are 80.6% and 81.6% respectively. The 1% higher CGE for Case 1 compared to the reference Shell gasifier IGCC case is primarily due to the extra 1% heat loss for the Shell gasifier in addition to the gasifier wall duty of 2% of the feed HHV.

The CGE for Cases 2 and 3 which are based on the coal/NG co-feed GTI HMB gasifiers are 78.0% and 87.8% respectively.

Case 2 has the lowest CGE of the four cases. The primary reason is that a net 15,000 lbmol/h of steam feed to the gasifier is unreacted and needs to be heated to the syngas temperature of 2,600°F. This requires additional heating duties from the coal/NG feed. This additional feed duty increases the feed HHV and lowers the CGE (equation 1).

At 87.8%, Case 3 has the highest CGE of the four cases. The primary reason is that heat is recovered from the hot, raw syngas through heat exchange with endothermic steam reforming of natural gas in an external steam reformer, enabling chemical energy to be returned as fuel to the gasifier. The cooled syngas exits the steam reformer at ~ 1,806°F compared to 2,600°F for the other cases.

8.3. Cost Results

Except for the costs of the HMB gasifier, the steam reformer and the high temperature heat exchangers, the Total Plant Cost (TPC) for Case 1, Case 2 and Case 3 GTI HMB gasifier based IGCC plants was determined by capital cost scaling following the 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 reference Case S1B. Once these have been established, the capital cost is estimated based on the revised capacity from the heat and material balances developed by Nexant.

The Total Overnight Cost (TOC) is then calculated by adding the owner's cost to the TPC.

The TOC for the three GTI HMB cases are lowered than the reference Shell S1B case as shown in Table 8-3. The lower TOCs are primarily due to the lower cost of the GTI HMB gasifier. Among the GTI HMB cases, Case 3 has the lowest TOC for the three GTI HMB Cases.

A comparison of the reference Shell S1B case and the GTI HMB cases TPC details are shown in Table 8-4. Analysis of the TPC cost details identified the following cost differences:

- The cost of the GTI HMB Case 1 gasifier system is ~ 47% lower than the reference Shell case.

- The costs of the GTI HMB Case 2 and Case 3 gasifier systems are ~ 43% and 40% respectively lower than the reference Shell case. Cases 2 and 3 gasifier system costs include natural gas compression and preheat, steam preheat and high temperature exchangers. Case 3 gasifier system cost also includes the steam reformer cost.

- The coal handling and feed systems and the ash handling systems for the natural gas co-feed cases (Cases 2 and 3) are ~ 30% to 35% lower than the reference Shell case due to lower coal feed rates.

- CO_2 compression and AGR costs are lower for GTI Cases 2 and 3 because of lower carbon/MMBtu of feed for the natural gas co-feed cases as discussed in sections 5 and 6. The costs are ~15 to 23% lower for CO_2 compression and drying and 7% lower for AGR.

- Case 2 generates more steam for the HRSG due to high temperature syngas cooling. It has the highest net power generation (510 MW vs 460 MW) versus the reference Shell case. The cost of the HRSG is ~15% higher than the reference case.

The costs of the GTI HMB gasifier for the three GTI cases are based on the gasifier sizes as shown in the conceptual layout in figures 4-2, 5-2 and 6-2. The gasifier sizes are estimated using GTI's gasifier dimensions and refractory thickness and the wall heat exchange tube requirements. Allowance for burners, steam drums and circulating pumps, and slag removal are included in the gasifier cost.

The cost of the steam reformer for Case 3 is based on the reformer size as shown in the conceptual layout in figure 6-3. The reformer dimensions are estimated from the tube heat exchange surface and catalyst volume requirements based on reforming duty. Incoloy is assumed for the high temperature reforming tube material of construction to provide additional contingency for the reformer cost. Traditional steam reformer tube materials of construction are HK40 or IN-519 which are high Cr and Ni alloys for high temperature service. The reformer vessel wall is assumed to be 316SS construction with 6" of refractory.

8.4. Cost of Electricity

The figure-of-merit metric used to evaluate overall financial performance is the cost of electricity (COE) for the IGCC plant. All costs are expressed in the "first-year-of-construction" year dollars, and the resulting COE is also expressed in "first-year-of-construction" year dollars.

The same financial modeling methodology is used for this study as per the NETL 1399 Baseline Study, which in turn is consistent with 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, fixed and variable operating costs, capacity factor and net power generation as shown in section 2.8.1.

The capital charge factor (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.

As shown in Table 8-3, the COEs are lower for the three GTI HMB cases than the reference Shell case due to the lower TPC/TOC. Among the three GTI HMB cases, the fuel costs became significant and impacted cases 2 and 3 in terms of annual operating costs. However, the COEs for the three GTI HMB cases are close and within 1.5% of each other.

When CO_2 TSM are included in the COE calculation, there is a 2% savings in the COE due to the lower carbon/MMBtu of feed for the co-feed cases.

8.5. Environmental Performance

Table 8-10. IGCC Environmental Targets

Pollutant	Environmental Target	NSPS Limit
NOx	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

Emissions	Control Technology
Sulfur Recovery	Claus Plant with Tail Gas Treatment / Elemental Sulfur
Particulate Control	Cyclone, Candle Filter, Scrubber, and AGR Absorber
Mercury Control	Carbon Bed
NO_x Control	MNQC (LNB) and N_2 Dilution

- Emissions of SO2 are extremely low (<0.0128 lb/MMBtu).
- Particulate emissions are the same for each case because it was a study assumption that the combination of cyclones and candle filters would meet the environmental target of 0.0071 lb PM/MMBtu
- NOx emissions were assumed to be 15 ppmv at 15 percent oxygen.
- Mercury emissions are significantly below the NSPS limit of 20 x 10-6 lb/MWh for IGCC systems. For the co-feed cases of Case 2 and Case 3, the emissions are even lower due to lower coal feed.

Appendix A Acronyms and Abbreviations

°F	Degree Fahrenheit
AGR	Acid Gas Removal
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
CW	Cooling Water
DBT	Dry Bulb Temperature
DOE	U.S. Department of Energy
DSP	Dry Solids Pump
EPC	Engineering, Procurement and Construction
EPRI	Electric Power Research Institute
EPRI	Electric Power Research Institute
FO	Fuel Oil
FOA	Funding Opportunity Announcement
ft	feet
GE	General Electric
h	Hour
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen Sulfide
Hg	Mercury
HGCU	Hot Gas Clean Up

HHV	Higher Heating Value
HMB	Hybrid Molten Bed
HP	High Pressure
HRSG	Heat Recovery Steam Generator
I & C	Instrumentation & Control
IGCC	Integrated Gasification Combined Cycle
IOU	Investor Owned Utility
kWe	Kilowatt electric
kWh	kilowatt hour
lb	Pound Mass
LH	Lock Hopper
LP	Low Pressure
max	Maximum
ME	Major Equipment
MEC	Major Equipment Cost
min	Minimum
Misc	Miscellaneous
MM	million
MP	Medium Pressure
MU	Makeup
MWe	Megawatt electric
MWh	megawatt hour
N_2	Nitrogen
NETL	National Energy Technology Laboratory
NOx	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
PFD	Process Flow Diagram
PM	Particulate Matter
ppmv	Parts per Million by Volume
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
SC	Supercritical
scf or SCF	Standard Cubic Feet
scfh or SCFH	Standard Cubic Feet per Hour
scfm or SCFM	Standard Cubic Feet per Minute
SO_2	Sulfur Dioxide
SOPO	Statement of Project Objectives
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	Short Tons per Day
US, USA	United States of America
vol%	Percentage by Volume
WBT	Wet Bulb Temperature
WGS	Water Gas Shift
WT	Waste Treatment
wt%	Percentage by Weight

Hybrid Molten-Bed Gasifier for Production of High Hydrogen Syngas

Techno-Economic Analysis Fischer-Tropsch Coal/Natural Gas to Liquid Production Option

Submitted to

GTI / DOE NETL

By

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1. INTRODUCTION

1.1. Background

Under the Department of Energy (DOE) Funding Opportunity Announcement (FOA) Number: DE-FOA-0000784, entitled "*Advanced Gasification Technologies Development and Gasification Scoping Studies for Innovative Initiatives*", Gas Technology Institute (GTI) is developing an innovative hybrid molten bed (HMB) gasification process to produce high-hydrogen syngas using coal and natural gas co-feeds and integrating with a nominal 50,000 barrels per day Fischer-Tropsch (FT) plant to produce diesel and naphtha liquid fuels. The coal/natural gas to liquid (CNTL) plant design will be based on a stand-alone greenfield facility located at Midwestern United States using Illinois No. 6 bituminous coal as the coal feed. This study will analyze the technology and economics of the GTI's hybrid molten bed (HMB) gasification process in conjunction with the Fischer-Tropsch process for liquid fuels production. The CNTL plant will also be design to limit the atmospheric carbon emission to less than 10% by capture at least 87% of the feed carbon content as FT products and CO_2.

1.2.Study Objectives

This techno-economic analysis (TEA) study was carried out to evaluate the hybrid molten bed (HMB) gasification process in the context of a Fischer-Tropsch CNTL production plant with CO_2 capture.

The objective of this techno-economic analysis is to assess the cost and performance of an FT CNTL plant with CO_2 capture that utilizes GTI's hybrid molten bed (HMB) gasification process to gasify Midwestern Illinois No. 6 coal. The GTI HMB gasifier is a dual coal-natural gas fueled molten bed gasification process. By varying coal and natural gas feed rates, and steam to natural gas ratio to the gasifier, the syngas H_2/CO ratio can be enhanced in the HMB gasifier to improve the overall FT CNTL plant efficiency and reduce the plant cost.

Three GTI HMB Gasifier cases with variations in feed mix, gasifier configuration and heat integration schemes are analyzed for the FT CNTL plant with CO_2 capture option. These cases are evaluated against a reference Shell SCGP gasifier based FT CNTL plant with CO_2 capture. Schematic depictions of these cases are included in the simplified block flow diagrams in figures 4-1 to 4-4 in section 4. The four cases studied are:

- Reference Case - Shell SCGP Gasifier FT CNTL Plant with 100% Illinois No. 6 Coal Feed and CO_2 Capture
- Case 1FT- GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture
- Case 2FT- GTI HMB Gasifier FT CNTL Plant with 81% Coal / 19% NG Feed and CO_2 Capture (Parallel Indirect Reforming)
- Case 3FT - GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture (Series Indirect Reforming)

The reference FT CNTL plant for the techno-economic analysis is based on the Shell gasifier with 100% Illinois No. 6 coal feed and with CO_2 capture. The FT section of the prototype is from Nexant's past FT estimates for iron-catalyst FT plants. Section 4 provides more details on the prototype design. It is designed to produce a nominal 50,000 BPD of FT diesel and naphtha.

Case 1FT is designed to take advantage of the dual feed capability of the GTI HMB gasifier design by using an optimum coal/NG feed mix to generate in the HMB gasifier the required H_2/CO ratio of 1.5 for the iron based FT synthesis. This will eliminate the need for water gas shift reactors and hence reduce the cost of the FT CNTL plant. The FT CNTL plant is designed to produce a nominal 50,000 BPD of FT diesel and naphtha.

Case 2FT is designed to generate the required H_2/CO ratio of 1.5 for the iron based FT synthesis using the parallel indirect reforming configuration. This configuration utilizes an external steam methane catalytic reformer where natural gas and/or FT tail gas is reformed with steam. However, instead of returning the reformer syngas to the HMB gasifier, the relatively clean reformer syngas is cooled and processed for contaminant removal separately from the gasifier syngas. The reformer duty is provided by the 2,600°F syngas exiting the GTI HMB gasifier.

Case 3FT is designed to generate the required H_2/CO ratio of 1.5 for the iron based FT synthesis using the series indirect reforming configuration. This configuration utilizes an external steam methane catalytic reformer where natural gas and/or FT tail gas is reformed with steam. The reformer duty is provided by the 2,600°F syngas exiting the GTI HMB gasifier. Case 3FT differs from Case 2FT in that the reformer syngas is returned to the HMB gasifier through the dual feed gasifier burners carrying with it the recuperated heat from the gasifier syngas.

The four cases are evaluated and compared based on their overall merits in terms of their cost of production (COP).

1.3. Plant Cost of Production Results

A summary of the FT CNTL plant cost of production results is shown in Table 1-1.

Table 1-1
FT CNTL Plant Cost of Production Summary

Case	Reference Shell Gasifier FT CTL	Case 1FT Direct Reforming FT CNTL	Case 2FT Parallel Indirect Reforming FT CNTL	Case 3FT Seriesl Indirect Reforming FT CNTL
2011 Capital Cost, $MM				
Total Plant Cost, $MM	6,543	5,702	5,571	5,742
Total Overnight Cost, $MM	8,078	7,116	6,920	7,139
2011 Operating Cost, $MM/yr				
Fixed Operating Costs	240	214	210	215
Variable Operating Costs @ 90% CF	186	147	152	141
Fuel Costs @ 90% CF, Coal @$68.6/ton	518.3	349.9	466.5	303.4
NG @ $5.17/MMBtu	0.0	506.1	191.7	428.6
Total Fuel Cost	518.3	856.0	658.2	732.0
COP FT Diesel, excl CO2 TS&M, $/bbl FT diesel	174	187	167	174
COP FT Naphtha, excl CO2 TS&M, $/bbl FT Naphtha	121	130	116	121
COP FT ECO, excl CO2 TS&M, $/bbl ECO	135	145	130	135
COP FT EPD, excl CO2 TS&M, $/bbl EPD	168	181	162	169

Of the four FT CNTL cases analyzed, Case 2FT is $7/Bbl FT diesel or 4% lower in COP relative to the reference Shell gasifier case. This case is configured with GTI HMB gasifiers and parallel indirect natural gas steam reforming and requires 81% coal and 19% natural gas as feed to produce 49.955 BPD of FT liquid fuels. Table 1-2 shows a comparison of the techno-economic performance for the four FT CNTL cases. The comparison identifies the key reasons for Case 2FT having the lowest COP.

Table 1-2
FT CNTL Plant Techno-Economic Performance Summary

Case	Reference Shell Gasifier FT CTL	Case 1FT Direct Reforming FT CNTL	Case 2FT Parallel Indirect Reforming FT CNTL	Case 3FT Seriesl Indirect Reforming FT CNTL
FT Liquid Fuels Products				
FT Diesel, BPD	38,053	36,611	37,193	37,335
FT Naphtha, BPD	13,057	12,562	12,762	12,811
Total FT Liquid Fuels, BPD	51,110	49,173	49,955	50,146
Feed Mix (HHV)				
Coal, MMBtu/hr	22,360	15,094	20,126	13,087
Natural Gas, MMBtu/hr	0	12,406	4,699	10,506
Total, MMBtu/hr	22,360	27,500	24,825	23,593
% Coal	100	55	81	55
% NG	0	45	19	45
Coal, TPD As Received	23,000	15,527	20,702	13,462
Oxygen Feed, TPD (100% O2 Basis)	18,508	27,465	17,019	18,404
FT Feed Gas				
H2/CO, mol/mol	1.47	1.49	1.48	1.49
CO2 to Sequestration, TPD	31,755	27,963	27,592	20,643
Power Production, Mwe				
Gas Turbine	260	190	414	271
Steam Turbine	407	579	161	303
Auxiliary Power Consumption	664	763	574	566
Net Power Output	3	6	1	8
Raw Water Withdrawal, gpm	47,729	32,072	34,280	25,608
No. of Gasifiers (including spares)	9	22	11	18
No. of Reformer	0	0	12	14
No. of ATR	4	0	0	0

The relative oxygen feed requirement is a very good indication of the gasification efficiency and gasification system cost when comparing gasification processes. Case 2FT requires the least amount of oxygen feed among the four cases. This is the result of gasifying coal only in the gasifier and reforming natural gas with steam external to the gasifier. In other cases, oxygen is used to gasify coal and also gasify and heat the co-feeds to the gasifier. Hence, more oxygen is required for the other cases. The following discussion compares Case 2FT which has the lowest oxygen feed to the other three FT CNTL cases:

- The reference Shell case has higher coal feed rate than Case 2FT (100% coal feed (23,000 TPD) vs 81% coal feed mix (20,702 TPD) for Case 2FT). More oxygen is required.

- Although Case 1FT uses less coal (55% coal mix, 15,527 TPD vs. 20,702 TPD), it also uses natural gas and steam as co-feed and the resulting syngas must also be heated to the gasification temperature of 2,600°F. Hence, more oxygen is required to gasify and heat the coal, natural gas and steam co-feeds to the gasifier for Case 1FT.

- Case 3FT uses the least coal (55% coal mix, 13,462 TPD) and also has the advantage of recycling the heat from the reformer syngas. It requires more oxygen than Case 2FT because of the need to also heat the recycled reformer syngas to 2,600°F. This is reflected in this case by having the second lowest oxygen requirement.

The oxygen requirement is also an indication of the amount of gasification syngas generated which impacts the size and cost of the gasification trains. The gasification train is consisted of the gasifiers, ASU, coal handling and conveying, natural gas compression, steam reformer, syngas heat recovery and syngas cleaning.

The number of GTI HMB gasifiers determines the size of the gasification train. The number was estimated based on the gasifier syngas rate and the residence time of 4 seconds per gasifier. It can be seen that Case 1FT has the highest number of HMB gasifiers (22) and case 2FT has the lowest number of HMB gasifiers (11). The impact of the gasification train on the TPC for the gasification system is shown in Table 1-3.

Table 1-3
FT CNTL Plant Gasifier System Cost Summary

TOTAL PLANT COST, 2011 $MM	Reference Shell Gasifier FT CTL	Case 1FT Direct Reforming FT CNTL	Case 2FT Parallel Indirect Reforming FT CNTL	Case 3FT Seriesl Indirect Reforming FT CNTL
GASIFIER & ACCESSORIES				
Gasifier, Quench Column, Filters & Cyclones	1,671.4	896.3	488.2	798.9
Steam Reformer	-	-	472.3	551.0
Natural Gas Compression	-	17.4	8.0	9.3
Syngas Heat Recovery	-	106.7	30.5	61.3
ASU/Oxidant Compression	673.4	789.0	619.1	653.9
LT Heat Recovery & FG Saturation	85.9	69.0	44.9	60.3
Flare Stack System	6.6	5.4	6.2	5.0
Gasification Foundations	93.8	77.1	89.0	71.8
Total	2,531.1	1,960.9	1,758.2	2,211.5

The lower number of gasifiers for Case 2FT corresponds to the lower cost of the HMB gasifiers and the gasification system TPC. It can be seen in Table 1-3 that the lower cost of the gasifiers for Case 2FT along with the lower cost of the ASU, natural gas compression and syngas heat recovery offset the additional cost of the reformers.

In conclusion, Case 2FT with parallel indirect reforming is recommended for further study and development because of its lower COP. Areas where further cost reductions are possible are in the further development of the gasifier and gas to gas steam methane reformer design.

2. FT CNTL DESIGN STUDY BASIS

2.1.Design References

The subject TEA study was carried out in accordance to DOE NETL's provided study guideline and the recommended reference studies set forth by Attachment 2 of the FOA that include the following:

NETL's *"Cost and Performance Baseline for Fossil Energy Plants Studies*" referred to as *"*Baseline Studies"[4] contained a comprehensive set of design basis and economic evaluation assumptions and criteria. These will be served as references for the purpose of the current study. DE-FOA-0000784 ATTACHMENT 2 also listed the following Baseline Studies references:

9. "*Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity* (Original Issue Date, May 2007), NETL Report No. 2010/1397, Revision 2, August 2010" - (NETL Report 1397)
10. "Cost and Performance Baseline for Fossil Energy Plants, Volume 4: Bituminous Coal to Liquid via Fischer-Tropsch Synthesis" (May 12, 2014) -------------- (NETL 2011/1477)
11. NETL's "Production of Zero Sulfur Diesel from Domestic Coal," referred to as the "CBTL study" (December, 2011) -- (NETL 2012/1542)

The following recommended QGESS reports are also used to provide consistent design basis for feedstock and equipment specifications, and cost estimation methodology:

12. "*Detailed Coal Specifications*, NETL Report No. 401/012111, January 2012" - (NETL Report 401/012111)
13. "*Process Modeling Design Parameters*, NETL Report No. 341/081911, January 2012" - (NETL Report 341/081911)
14. "*Specification for Selected Feedstocks*, NETL Report No. 341/011812, January 2012" - (NETL Report 341/011812)
15. "*CO_2 Impurity Design Parameters*, NETL Report No. 341/011212, August 2013" - (NETL Report 341/011212)

NETL Report 1399 provides reference costs and economic evaluation guidelines. Additionally, the following reports also serve as reference sources for the economic evaluation reference in this study.

16. "*Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases*, August 2012, DOE/NETL-341/082312"- (NETL Report 341/082312)
17. NETL's Series of Quality Guidelines for Energy Systems Studies (QGESS):
    - "*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"
    - "*Fuel Prices for Selected Feedstocks in NETL Studies*, November 2012, DOE/NETL 341/11212"

[4] http://www.netl.doe.gov/energy-analyses/baseline_studies.html

2.2.Process Design Parameters

2.2.1. Coal Properties and Firing Rate

Design coal feed to the FT CNTL production plants is Illinois No. 6 bituminous coal with characteristics presented in Table 2-1. The as-received coal properties shown in Table 2-1 are from the QGESS *Detailed Coal Specifications* document. The as-received coal is dried to 5% moisture and fed through to the Shell or GTI HMB gasifier. The gasifier will gasify enough dried Illinois No. 6 coal to produce sufficient syngas for a nominal 50,000 barrels per day Fischer-Tropsch plant to produce diesel and naphtha liquid fuels.

Table 2-1
Illinois No. 6 Coal Specification

Rank	Bituminous	
Seam	Illinois #6 (Herrin)	
Sample Location	Old Ben Mine	
Ultimate Analysis, weight%	As-Received	Dry
Moisture	11.12	0.00
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
Ash	9.70	10.91
Oxygen (by difference)	6.88	7.75
Total	100.0	100.0
Proximate Analysis[a], weight%	As-Received	Dry
Moisture	11.12	0.00
Ash	9.70	10.91
Volatile Matter	34.99	39.37
Fixed Carbon (by difference)	44.19	49.72
Total	100.0	100.0
Higher Heating Value (HHV), Btu/lb	11.666	13,126
Sulfur Analysis[b], 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 1399 Baseline Study, this study assumes that all sulfur in the coal is converted in the gasifier and leaves with the syngas

2.2.2. Natural Gas Properties

The design composition for the natural gas feed to the FT CNTL production plant is shown in Table 2-2.

Table 2-2
Natural Gas Composition & Heating Values

Component	Volume Percentage	
Methane, CH_4	93.1	
Ethane, C_2H_6	3.2	
Propane, C_3H_8	0.7	
n-Butane, C_4H_{10}	0.4	
Carbon Dioxide, CO_2	1.0	
Nitrogen, N_2	1.6	
Total	100.0	
	LHV	HHV
Btu/SCF	932	1,032
Btu/lb	20,410	22,600

2.2.3. Gasification Block Process Design Criteria

GTI Block is designed as an integral part of the FT CNTL plant. It includes the following major gasification and syngas cleanup related systems:

- Feed Pressurization and Drying System
- HMB Gasifier
- Steam Methane Reformer (if used)
- Syngas Cooling and Reforming Steam Generation
- Rectisol AGR
- CO_2 Compression and Purification Facilities
- O_2 Booster Compressor
- NG Booster Compressor and Preheat
- Reforming Steam Preheat

The process design parameters for the GTI gasification block are summarized in Table 2-3. The reference case gasification island design parameters are based on DOE's baseline study report DOE/NETL 2011/1477.

Table 2-3
Gasification Block Process Design Parameters

Case	Reference Case	GTI HMB Gasifier Cases
Gasifier Technology	Shell (SCGP)	GTI (HMB)
Coal Energy Content (%)	100%	>50%
Gasifier Pressure, (psia)	545	As Required
O_2:Coal Ratio, lb O_2/lb dry coal	0. 86	As Required
Carbon Conversion, %	99.5	by GTI
Gasifier Heat Removal by Steam Generation, % Feed HHV	2%	2%
Gasifier Heat Loss, % of Feed HHV	1%	0%
Nominal Steam Cycle, (psig/°F/°F)	1,800/1,000/1,000	1,800/1,000/1,000
Condenser Pressure, (in Hg)	1.4	As Required
Combustion Turbine	GE SG6FA	GE SG6FA
Oxidant	95 vol% Oxygen	Same
Coal	bituminous	Same
H_2S Separation	Rectisol	Same
Sulfur Removal, %	99.7	As Required
CO_2 Separation	Rectisol	Same
CO_2 Emission, %	<10%	<10%
Sulfur Recovery	Claus Plant with Tail Gas Treatment / Elemental Sulfur	Same
Particulate Control	Cyclone, Candle Filter, Scrubber, and AGR Absorber	Same
Mercury Control	Carbon Bed	Same
NO_x Control	MNQC (LNB) and N_2 Dilution	Same

2.2.4. FT Liquid Fuels Production Block Design Criteria

The FT liquid fuels production block referenced design is based on Nexant's in-house FT data. It was developed by Nexant for use in the techno-economic study of the GTI HMB gasifier plant with CO_2 capture. The design criteria are shown in Table 2-4. The GTI HMB CNTL plant FT Block costs will be scaled against the reference FT block cost based on capacity factors.

Table 2-4
FT Liquid Fuels Production Block Process Design Parameters

	DOE.NETL-2011/1477	Reference Shell Gasifier Case
As-received Coal Feed Rate, ton/day	21,006	TBD
H_2/CO, (lbmol/lbmol)	0.73	1.5
FT Feed Pressure, psia at inlet to FT reactors	325	413
Diesel + Naphtha Production, bbl/day	50,000	TBD

2.2.5. Non-GTI Block Design and Criteria

The Non-GTI Blocks (NGB) include the systems common to both the conventional and GTI HMB FT CNTL plants, which are not directly related to the advanced coal gasification, syngas cleanup and FT liquid production systems. Apart from being of different capacities, these systems are expected to have nearly identical flow schemes as the corresponding conventional FT CNTL with CO_2 capture reference case. Due to the similarity in designs between these systems that are common to both the advanced and conventional FT CNTL cases, the Non-GTI Block systems costs will be scaled based on capacity factors given in the QGESS *Capital Cost Scaling Methodology* document for the advanced IGCC plant wherever possible.

Process modeling for the NGB systems will be carried out, to the maximum extent possible, in accordance with guidelines from the Baseline Study NETL 1477 and QGESS *Process Modeling Design Parameters* documents. This is used mainly to determine the utilities consumption or power generation rates of the NGB systems in order to evaluate the overall FT CNTL plant efficiency.

2.2.6. Turbine Design Criteria

The system power for FT CNTL plant is supplied by the gas turbine (GT) and the steam turbine ST) combined cycle plant. The reference GT was selected based on the largest commercially available syngas-fired GT with fuel pressure requirement closely matches the FT plant purge gas pressure. The General Electric SG6FA gas turbines using FT purge (tail-gas) to generate power was used based on GT supplier quotes from past CTL projects using Nexant's FT plant design. At ISO condition, each SG6FA GT gross power output is roughly 95 MWe, as measured prior to the generator terminals.

2.2.7. Steam Cycle Design Criteria

The power plant is also equipped with a heat recovery steam generator (HRSG) coupled with steam turbine to recover waste heat from the GT flue gas to generate additional power. Total power generated by the GT and ST will supply the total plant auxiliary load with minimal excess (near zero) for power export.

The HRSG is a horizontal gas flow, drum-type, multi-pressure design that is matched to the characteristics of the gas turbine exhaust gas. The HRSG/steam turbine power cycle will be modeled based on the assumed ambient conditions, back-end loss, and HRSG pressure drop. The ST cycle consists of three pressure levels: 1500 psig (HP), 150 psig (MP), and 50 psig (LP). Saturated high pressure (HP) steam mainly from Gasification and HRSG are superheated in the

HRSG to about 950°F before entering the HP stage of the ST. Intermediate pressure (IP) steam at 650 psig and 300 psig are extracted from the HP ST to meet process demands. Exhaust from the HP ST is mixed with the saturated medium pressure (MP) steam generated in the HRSG and FT plant before entering the MP stage of the ST without reheat in the HRSG. Exhaust from the MP ST is mixed with the saturated low pressure (LP) steam generated in the HRSG before entering the LP ST. Exhaust from the LP ST is condensed at 4" Hg (2 psia) via water cooled surface condensers. The main steam conditions are shown in Table 2-5.

Table 2-5
Steam Conditions for the CNTL Plant

Main Steam Pressure, psig	1,800
Main Steam Temperature, °F	1,000 (Range 950-1075)
Reheat Steam Temperature, °F	1,000 (Range 950-1075)

2.2.8. Cooling Water

The CNTL plant cooling water system is based on the guidelines as described in the NETL QGESS titled "Process Modeling Design Parameters, Rev. January 17, 2012". A mechanical draft, evaporative recirculating wet cooling tower is used, and all process blowdown streams are assumed to be treated and recycled to the cooling tower. Typical cooling tower approach temperatures are in the range of 8 to 20°F for the power plant applications. NETL systems studies use an approach to wet bulb of 8.5°F for ISO location. The design ambient wet bulb temperature of 51.5°F is set to achieve a cooling water temperature of 60°F using an approach of 8.5°F. Cooling water range is assumed to be 20°F. Cooling water from the cooling towers is available at the following conditions:

- Maximum supply temperature, °F 60
- Maximum return temperature, °F 80

Cooling tower makeup rate calculation is also specified by the same NETL QGESS, and is determined as followed:

- Evaporative losses = 0.8 percent of the circulating water flow rate per 10°F of range
- Drift losses = 0.001 percent of the circulating water flow rate
- Blowdown losses = Evaporative Losses / (Cycles of Concentration - 1)
  where cycles of concentration are a measure of water quality and a mid-range value of 4 is chosen for this study

2.2.9. Air Separation Unit (ASU) Design Criteria

The air separation plant is designed to produce 95 mole percent O_2 for use in the gasifier. The air compressor is powered by an electric motor. Nitrogen is also recovered, compressed, and used for fuel gas dilution in the GT combustor.

Conventional cryogenic ASU will be used to produce the 95 mole percent purity oxygen for use in the GTI HMB gasification. The ASU will be designed for ambient air quality as shown in Table 2-6. Product oxygen composition is listed in Table 2-7 below. An oxygen compressor will be provided to boost the product oxygen pressure to that required to feed the GTI HMB gasifier.

ASU performance and utility consumption will be pro-rated from the NETL Report 1477 design based on total oxygen production.

Table 2-6
Ambient Air Quality

Air composition based on published psychrometric data, mass %	
Argon	1.283
CO_2	0.050
O_2	23.049
N_2	75.220
Moisture	0.398
Total	100.00
Air Composition, mol%	
Argon	0.93
CO_2	0.03
O_2	20.81
N_2	77.59
Moisture	0.64
Total	100.00
Site Conditions:	
Ambient Pressure, psia	14.7
Design Ambient Temperature, Dry Bulb, °F	42
Design Ambient Temperature, Wet Bulb, °F	37
Design Ambient Relative Humidity, %	62

Table 2-7
Product Oxygen Quality

Analysis by Weight:	Volume %
N_2	1.78
O_2	95.04
Argon	3.18
Total Vol%	100.00
Conditions before Booster Compression:	
Pressure, psia	125
Temperature, °F	90

2.2.10. Balance of Plant

Balance of Plant design basis such as fuel and chemical storage and plant distribution voltages are summarized in Table 2-8.

Table 2-8
Balance of Plant

Fuel and Other Storage	
Coal	30 days
Slag	30 days
Sulfur	30 days
Sorbent	30 days
Plant Distribution Voltage	
Motors below 1 hp	110/220 volt
Motors between 1 hp and 250 hp	480 volt
Motors between 250 hp and 5,000 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam and GT Generators	24,000 volt
Grid Interconnection Voltage	345 kV

2.2.11. CO_2 Product Treating and Purification Design Criteria

For this study, recovered CO_2 is delivered at the battery limit (B/L), with specifications for saline reservoir sequestration as listed in Table 2-9, per the NETL "CO_2 Impurities Design Parameters, Draft Report, August 23, 2013" QGESS reference. The one exception is that the CO content in the recovered CO_2 will be around 2500 ppmv for this study. The high CO concentration is estimated by the licensor for a Rectisol design modified for maximum CO_2 recovery in order to meet < 10% carbon emission requirement.

Table 2-9
B/L CO_2 Pipeline Specifications[5,6]

B/L Pipeline Pressure, psia	2,215
B/L Pipeline Temperature, °F	95
Compositions:	
CO_2, vol% (Min)	95
N_2 + Ar, vol% (Max)	4
O_2, vol% (Max)	4
CH_4 + H_2, vol% (Max)	4
CO, ppmv (Max)	2500
SO_2, ppmv (Max)	100
NO_x, ppmv (Max)	100
H_2O, ppmv (Max)	300

CO_2 compression facilities will be provided to boost the CO_2 product pressure to the required B/L requirement.

[5] http://www.netl.doe.gov/energy-analyses/pubs/LR_IGCC_FR_20110511.pdf
[6] http://www.netl.doe.gov/energy-analyses/refshelf/PubDetails.aspx?Action=View&PubId=420

2.2.12. Water Supply and Waste Water

Makeup Water
The water supply is 50 percent from a local publicly owned treatment works (POTW) and 50 percent from groundwater, and is assumed to be in sufficient quantities to meet plant makeup requirements. Makeup for potable, process, and de-ionized (DI) water is drawn from municipal sources.

Process Wastewater
Water associated with gasification activity and storm water that contacts equipment surfaces is collected and treated for discharge through a permitted discharge.

Sanitary Waste Disposal
Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site. Packaged plant was sized for 5.68 cubic meters per day (1,500 gallons per day)

Water Discharge
Most of the process wastewater is recycled to the cooling tower basin. Blowdown is treated for chloride and metals, and discharged.

2.2.13. Environmental/Emissions Requirements

The FT CNTL plant is a gasification/synfuels refining complex. The environment targets for this study were established in the Electric Power Research Institute's (EPRI) design basis for their CoalFleet for Tomorrow Initiative, documented in the CoalFleet User Design Basis Specification for Coal-Based Integrated Gasification Combined Cycle (IGCC) Power Plants, EPRI, Palo Alto, CA, 2009. The design targets were established specifically for bituminous coal but apply to subbituminous case as well. The emissions requirements and limits for the reference FT CNTL plant, as specified in NETL Report 1399, are listed in Table 2-10:

Table 2-10
FT CNTL Environmental Targets

Pollutant	Environmental Target	NSPS Limit
NOx	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 for the GTI HMB gasification FT CNTL process.

2.2.14. Overland Transportation Size Limitations

The site is landlocked with access by train and highway only. Maximum overland highway transportable dimension is assumed to be 100 feet long by 12 feet wide by 15 feet height (including carriage height). Maximum equipment height is 13.5 feet assuming using 1.5 feet height low boy carriage. Maximum overland highway transportable weight is 65 tons.

Maximum railway transportable dimension is assumed to be 100 feet long by 12 feet wide by 19 feet height (including railcar height). Maximum equipment height is 15 feet assuming using 4 feet height railcar. Maximum railway transportable weight is assumed to be 130 tons.

2.2.15. 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

2.3. Site-Related Conditions

The FT CNTL plants in this study are assumed to be located in Midwestern United States, with site-related conditions as shown below:

Location	Midwestern, US
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 2,200 psig at IGCC battery limit and transported 50 miles for sequestration in a saline formation at a depth of 4,055 ft (Study scope limited to delivery at battery limit only)

2.4. 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), °F 59
- Wet bulb temperature (WBT), °F 51.5
- Ambient relative humidity, % 60

2.5. Capital Cost Estimation Methodology

2.5.1. General

For the FT CTL plants with CO_2 capture, the NETL 1477 Baseline 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 studied GTI HMB gasifier based FT CNTL plant designs, except for the costs of the HMB gasifier, the steam methane reformer, and the FT units, capital costs were estimated by capacity scaling in according to the 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 Baseline Study. Once these have been established, the capital cost can be estimated based on the revised capacity from the heat and material balances developed for the design.

As defined in the DOE 1477 report, an average labor wage at $39.7/hour, with an all-in labor cost of $51.6/hour (including wages plus 30% burden to cover fringe benefits, payroll based taxes, and insurance premiums) is assumed for calculating the 2011 installation labor costs. No over-time or other premiums is added. The average labor productivity for the site is assumed to be 105% of the US Gulf coast productivity.

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

Construction indirect cost are 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 1477 report.

2.5.2. Balance of Plant Capital Cost Estimate Criteria

For the rest of the systems that are not related to coal handling, the capital cost estimates are developed based on the reference CTL plant with CO_2 capture case in NETL 1477 Baseline Study.

For these subsystems, capital cost scaling following the guidelines and parameters described in the NETL *Capital Cost Scaling Methodology* document is used to perform the cost estimates, as described in Section 2.5.1.

Table 2-11 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 BOP subsystems, as laid out by the NETL *Capital Cost Scaling Methodology* document, are also shown in this table.

Table 2-11
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, Quench Column, Filters & Cyclones	Calculated & Syngas Throughput
4.1a	Steam methane reformer	Calculated
4.1b	Natural Gas Compression	Calculated
4.2	Syngas Heat Recovery	Calculated
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
Acct No.	**Item/Description**	**Scaling Parameter**
5A	GAS CLEANUP & PIPING	
5A.1	Rectisol	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	Gas Turbine Power
6.2	Combustion Turbine Foundations	Gas Turbine Power
7	HRSG, DUCTING & STACK	
7.1	Heat Recovery Steam Generator	Steam Turbine Power
7.3	Ductwork	Steam Turbine Power
7.4	Stack	Steam Turbine Power
7.9	HRSG, Duct & Stack Foundations	Steam Turbine Power
8	STEAM TURBINE GENERATOR	
8.1	Steam TG & Accessories	Steam Turbine Power
8.2	Turbine Plant Auxiliaries	Steam Turbine Power
8.3a	Condenser & Auxiliaries	Steam Turbine Power
8.3b	Air Cooled Condenser	Steam Turbine Power
8.4	Steam Piping	Steam Turbine Power
8.9	TG Foundations	Steam Turbine Power
9	COOLING WATER SYSTEM	
9.1	Cooling Towers	Cooling Tower Duty
9.2	Circulating Water Pumps	Circ H_2O Flow Rate
9.3	Circ Water System Auxiliaries	Circ H_2O Flow Rate
9.4	Circ Water Piping	Circ H_2O Flow Rate
9.5	Makeup Water System	Raw Water Makeup
9.6	Component Cooling Water System	Circ H_2O Flow Rate
9.9	Circ Water System Foundations	Circ H_2O 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

Acct No.	**Item/Description**	**Scaling Parameter**
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 Pump House	Circ H_2O 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

2.5.3. Home Office, Engineering Fees and Project/Process Contingencies

Engineering and Construction Management Fees and Home Office cost, project and process contingencies will be factored from the each subsystem's TFC. These are then added to the TFC to come up with the total project cost (TPC) of the system. Factors from the NETL 1477 Baseline Report will be used for this study.

2.5.4. Owner's Cost

Owner's cost is then added to TPC to come up with the total overnight cost (TOC) for the system. Owner's costs as defined in the NETL 1477 Baseline 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

- Initial Cost for Catalyst and Chemicals per design
  - Land Cost = $900,000 at 300 acres x $3,000/acre
  - Other Owner's Costs at 15% TPC
  - Financing Costs at 2.7% TPC

2.6. Operation & Maintenance Costs

The operation and maintenance (O&M) costs pertain to those charges associated with operating and maintaining the CNTL 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 are estimated based on 90% capacity factor.

2.6.1. Fixed Costs

Operating labor cost is 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, $/hour	$39.7
Length of work-week, hours	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

2.6.2. Variable Costs

The cost of consumables, including fuel, is 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 are evaluated similarly to the consumables.

The unit costs for major consumables and waste disposal will be selected from NETL 1477 Baseline Report, 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 US power plant is $68.6/ton, per the QGESS *Fuel Prices for Selected Feedstocks in NETL Studies* document. The price of natural gas is $5.34/1000ft^3 ($5.17/MMBtu HHV) per QGESS.

2.6.3. CO_2 Transport and Storage Costs

As specified in DE-FOA-0000784 Attachment 2, CO_2 Transport and Storage (T&S) costs storage for the Midwestern FT CNTL plant location is $22/tonne. Per the TEA reporting requirements, the COPs will be reported both with and without the cost of CO_2 T&S.

2.7. Financial Modeling Basis

2.7.1. Cost of Production

The key measure to evaluate overall economic financial viability of the FT CNTL plant is the estimation of the crude oil equivalent required selling price (RSP) of the Fischer-Tropsch liquid products. The RSP is the minimum price at which the products must be sold to recover the annual revenue requirement (ARR) of the plant. The ARR is the annual revenue needed to pay the operating costs, service the debt, and provide the expected rate of return for the investors. The FT CNTL project is considered economic viable if the market price of the product is equal to or above the calculated RSP.

$$\text{RSP} = \frac{\begin{array}{c}\text{first year}\\\text{capital charge}\end{array} + \begin{array}{c}\text{first year}\\\text{fixed operating}\\\text{cost}\end{array} + \begin{array}{c}\text{fist year}\\\text{variable operating}\\\text{cost}\end{array}}{\text{annual net Fischer Tropsch Liquid production}}$$

$$\text{RSP} = \frac{(\text{CCF})(\text{TOC}) + \text{OC}_{\text{fix}} + (\text{CF})(\text{OC}_{\text{var}})}{\text{annual net Fischer Tropsch Liquid production}}$$

The ARR is the sum of fuel cost, variable operating cost, fixed operating cost, and annual capital component minus the by-product credits for electric power sale revenues. The annual capital component of the ARR is determined as the product of the total overnight cost (TOC) and the capital charge factor (CCF). The CCF for evaluating the RSP is determined from NETL Power Systems Financial Model (PSFM) using the financial parameters shown in Table 2-12. The capital charge factor of 0.218 for commercial fuels and 0.170 for loan guarantees will be estimated for the RSP (Equivalent Crude Oil) financial analysis.

Table 2-12
Code of Accounts for Report IGCC Plant

Scenario	Commercial Fuels	Loan Guarantees
Percent Debt	50%	60%
Percent Equity	50%	40%
Debt Interest Rate	8.00%	4.56%
Internal Rate of Return on Equity (IRROE)	20%	20%
After Tax Weighted Cost of Capital	12.48%	9.70%
Capital Charge Factor (CCF)	0.218	0.170

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 dollars.

The conceptual plants produce three products for sale. Those products are: (1) FT diesel fuel, (2) FT naphtha, and (3) electric power. All light gases including LPG are used within the plant. FT naphtha, although it has a similar boiling range to gasoline, has not traditionally been considered to be suitable for refining into high octane gasoline because of its high paraffinic nature. This analysis assumes that the naphtha can be sold at a discounted price compared to the diesel fuel. To express the RSP in terms of equivalent crude oil price (COP), historically, the ratio of the price of crude oil: ultra-low sulfur diesel is 1.25 and naphtha: diesel is 0.7. The discount price is assumed to be 0.7692 (1/1.3) the value of the diesel fuel. The relative value is used to determine the equivalent diesel fuel yield from the CTL plant in terms of barrels per year.

The petroleum equivalent diesel price is calculated by taking the first year of production for diesel in $/bbl and multiplying this value by the ratio of the lower heating values of FT diesel and petroleum diesel.

$$Petroleum\ Equivalent\ Diesel\ Price = \left(\frac{Petroleum\ Diesel\ LHV}{FT\ Diesel\ LHV}\right) * FY\ COP\ FT\ Diesel$$

- The equivalent crude oil price is then calculated by multiplying the petroleum equivalent diesel price by a factor of 0.80.

- RSP Equivalent Crude Oil = 0.80 x Petroleum Equivalent Diesel Price.

- 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. This data was used to develop correlations between the various fuel prices and the WTI crude oil price (Crude oil: Ultra-low sulfur diesel is 1.25 and Naphtha: Diesel is 0.70). The ECO price is the minimum market price for crude oil at which the first-year RSPs will be met.

Sensitivity analyses of FT liquids products required selling price (RSP) will be performed on the following parameters:

- Critical advanced technology performance parameters
- Capital cost of advanced technology
- Non-coal fuel prices
- Sales of CO_2 at plant gate prices of $0-60/tonne
- Cost of CO_2 emissions of $0-60/tonne
- Power price for net imports/exports at $60-135/MWh
- Finance structure by assessing capital charge factors of 0.12-0.25

The economic assumptions and finance structure/capital expenditure period is defined under High-Risk Fuels Projects in report DOE/NETL-2011/1489 September 29, 2011, (revision from DOE/NETL-401/090808) "*Recommended Project Finance Structure for the Economic Analysis of Fossil-Based Energy Project*". Listed below the financial parameters and assumptions required by the PSFM model: NETL/DOE-2011/1477 show repayment term of 30 years)

- Income tax rate, % 38
- Equity desired rate of return, % 20
- Type of debt financing Non-Recourse
- Repayment term of debt, years 30
- Debt repayment grace period, years 0
- Debt reserve fund None
- Depreciation 20 years, 150% declining balance
- Working capital None
- Plant operational life, years 30
- Plant economic life, years 35
- Tax holiday, years 0
- EPC escalation, % per year 3.6
- COE (revenue) nominal escalation, % 3.0
- Coal price nominal escalation, % 3.0
- O&M cost nominal escalation, % 3.0
- Duration of construction, years 5
- First year of construction 2011
- Construction cost distribution, %
  - Year 1 10%
  - Year 2 30%
  - Year 3 25%
  - Year 4 20%
  - Year 5 15%

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 dollars.

2.7.2. CO_2 Sales Price

As outlined in the TEA's reporting requirements, sensitivity analysis is to be done to determine the impact of CO_2 sales on FT CNTL COP. The varying parameter is the CO_2 sales price at the FT CNTL plant gate and is to range between $0/tonne (baseline case assuming no value to the product CO2) and $60/tonne.

Per the reporting requirements for the TEA, the cost of capturing CO_2 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, CO_2 avoided cost analysis for like technology reference will not be performed.

2.7.3. Cost of CO_2 Emissions

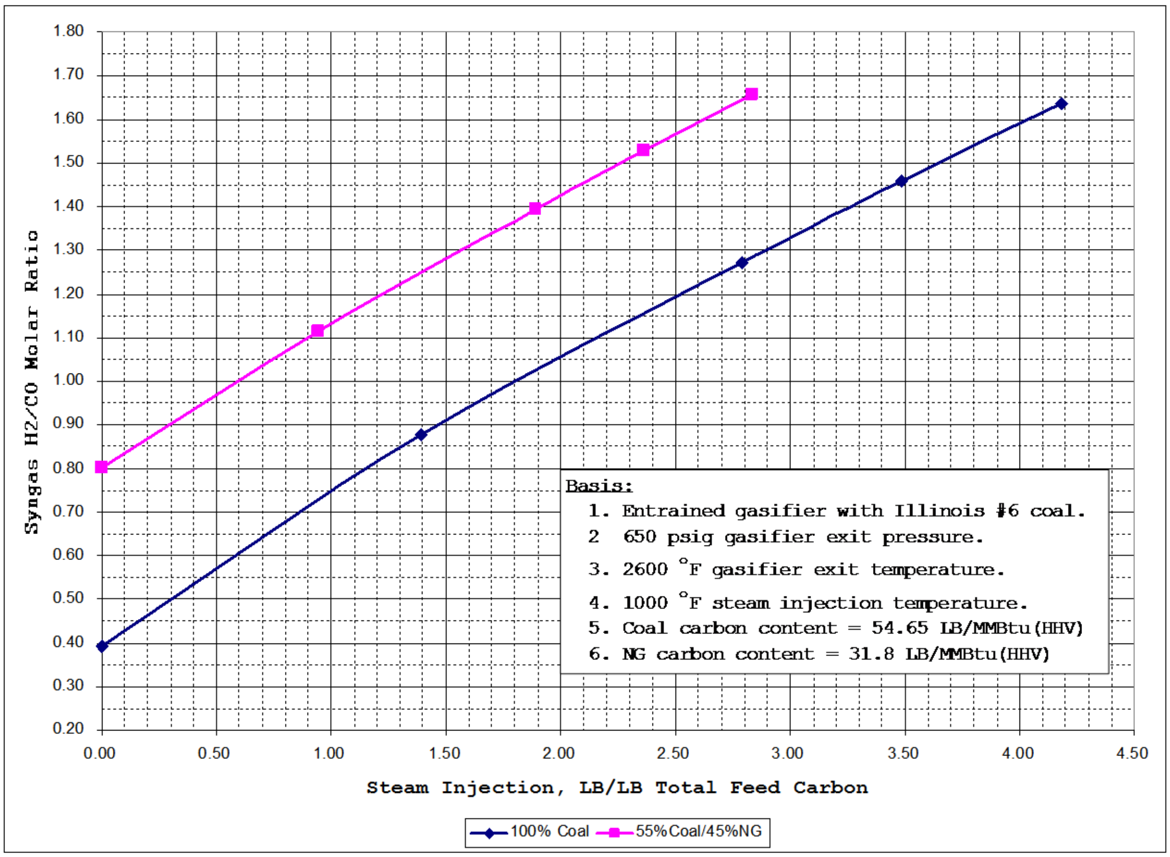
The TEA also requires sensitivity analysis on cost of CO_2 emissions to be performed. The varying parameter is the CO_2 emissions cost. The range of the emissions cost is between $0/tonne (baseline case assuming no CO_2 emissions cost) and $60/tonne.

3. GTI HYBRID MOLTEN BED GASIFIER

This study investigates the feasibility of generating syngas with H_2/CO ratio of 1.5, as required to feed commercially operating CTL FT plants based on iron catalyst technology. By generating 1.5 H_2/CO ratio syngas directly from gasification eliminates the need for separate downstream water-gas shift requirement. While it is theoretically possible to generate syngas with 1.5 H_2/CO ratio under entrained gasification conditions with steam injection, it will required steam-to-carbon mass ratio over 3.5-to-1 (lower curve in Figure 3-1), which is equivalent to a steam-to-coal mass ratio of 2.5-to-1 for Illinois #6 coal feed.

Since the NETL TEA guideline for this study allows co-feed of up to 45% of a second feed in combination with coal, steam injection versus syngas H_2/CO ratio is estimated for feed made up of 45% natural gas (NG) and 55% Illinois #6 coal, and is shown as the upper curve in Figure 3-1. The steam-to-carbon ratio required to generate syngas with 1.5 H_2/CO ratio is reduced to 2.3-to-1 when co-feeding NG, compared to the 3.5-to-1 needed for 100% coal feed.

Figure 3-1
Gasifier Outlet Syngas H_2/CO Ratio vs. Steam Injection



Currently no commercial dry feed entrained gasifier operates with steam-to-carbon ratio significantly above 1. Wet slurry-fed gasifier typically operates at feed moisture content of 35% for bituminous coal, and may be up to 50% for high moisture low rank coals. The corresponding water-to-carbon ratios are only about 0.7-to-1 for bituminous coal and 1.6-to-1 for low rank coals. These are much lower than that required to achieve 1.5 H_2/CO ratio with 100% coal feed.

Also, no commercial entrained gasifier can take advantage of the high NG co-feed opportunity to lower steam injection requirements since none are designed to operate with significant amount of NG co-feed together with coal.

The proposed GTI HMB gasifier is designed for dual feed and thus will be able to take advantage of the high NG co-feed opportunity. The ability to co-feed large amount of NG also provides the opportunity to recuperate some of the waste heat from the hot gasifier exhaust to provide the energy required for steam/methane reforming of the NG co-feed for conversion into H_2 and CO, in lieu of partial oxidation (POX) conversion by oxygen. Recycling of the syngas sensible heat should reduce the total amount of oxygen consumed for syngas generation and hence may improve the overall FT CNTL plant efficiency and cost. Various process schemes incorporating the GTI HMB gasifier to produce FT naphtha and diesel from coal plus up to 45% NG are evaluated against a referenced FT CTL plant based on Shell gasification of Illinois #6 coal. The Shell gasifier syngas H_2/CO ratio of approximately 0.45 is enhanced by water gas shift to a ratio of 1.5 before feeding to the FT synthesis system.

The simplest way to recuperate energy from the hot gasifier syngas would be by direct reforming. For direct reforming, the steam/natural gas (NG) second feed is mixed directly with hot HMB gasifier syngas in a separate reforming chamber. The total steam/methane/syngas mixture then reaches reforming equilibrium, either thermally at high temperature or catalytically at lower temperature. This is similar to commercial natural gas partial oxidation (POX) or autothermal reforming (ATR) processes but without burner firing. Current commercial NG POX processes are typically carried out in reaction temperature above 2,300°F in order to achieve reforming equilibrium without the use of catalysts. Commercial NG ATR processes are typically carried out in reaction temperature around 1,800°F with catalysts in order to achieve reforming equilibrium. As discussed in the HMB for IGCC application report, the minimum Coal-to-NG ratio needed to reach an adiabatic reaction temperature of 2,300°F without the aid of catalyst is 95% coal/5% NG (with steam-to-carbon ratio of 0.76), while that for adiabatic reaction temperature of 1,800°F with the aid of catalyst is increased to 85% coal/15% NG. However there is no commercial sulfur tolerant reforming catalysts available so the catalytic direct reforming option is not viable due to sulfur in the gasifier syngas. Thus, the coal-to-NG ratio is limited to 95% coal/5% NG for the non-catalytic direct reforming option, which is not much of an advantage over 100% coal feed operations especially when adjusted to the higher steam-to-carbon ratio (near 3.5) needed to get syngas with 1.5 H_2/CO ratio. Therefore, instead of direct reforming, the NG co-feed together with the required steam injection will be gasified together with the coal feed in the HMB gasifier for syngas generation.

Alternatively, an external reformer in parallel to the HMB gasifiers can be used to indirectly recuperate energy from the hot syngas. For indirect reforming, the steam/NG second feed goes through catalyst-packed high-alloy tubes and is heated indirectly by the external hot HMB syngas, similar to commercial convective reforming. The reformate product is then mixed with the cooled syngas before downstream upgrading processes. Similar to direct reforming, the indirect external reformer is also heated by syngas and thus its capacity for NG reforming is also affected by the coal/NG feed ratio. At fixed HMB gasification temperature, low coal/NG ratio means less hot syngas from coal gasification is available to heat the indirect reformer. Since it is operating at temperature range close to that for the direct reforming with ATR scheme, the minimum coal/NG limit for parallel indirect reforming is expected to be similar to that for direct reforming with ATR, about 85% coal and 15% NG.

To eliminate the minimum coal/NG restriction for the indirect reforming, the external reformer product can be routed back into the HMB gasifier to increase the syngas flow through the external reformer hot side. This will allow operation with coal/NG feed ratio of 55%/45%. This is the in-series indirect reforming scheme.

Three final FT CNTL configurations based on GTI recuperative HMB Gasification schemes were evaluated against the reference case developed by Nexant. The reference case is a Shell gasifier based FT CTL plant, designed with Nexant's in-house data for iron catalyst with slurry bed reactor-based commercial FT technology. The plant will produce 50,000 BPD of FT diesel and naphtha. Schematic depictions of these FT CNTL cases are included in the simplified block flow diagrams in figures 4-1 to 4-4 in section 4. The four cases studied are:

- Reference Case - Shell SCGP Gasifier FT CTL Plant with 100% Illinois No. 6 Coal Feed and CO2 Capture
- Case 1FT- GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture
- Case 2FT- GTI HMB Gasifier FT CNTL Plant with 81% Coal / 19% NG Feed and CO_2 Capture (Parallel Indirect Reforming)
- Case 3FT - GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture (Series Indirect Reforming)

4. FT CNTL CASE CONFIGURATIONS

4.1. FT CNTL Case Configurations

The techno-economics of three different HMB gasification FT CNTL configurations were studied, against the reference which is based on a Shell SCGP gasifier design. Schematic depictions of these cases are included in the simplified block flow diagrams of figures 4-1 to 4-4. The four case studies are as follows:

- Reference Case - Shell SCGP Gasifier FT CNTL Plant with 100% Illinois No. 6 Coal Feed and CO2 Capture
- Case 1FT- GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture
- Case 2FT- GTI HMB Gasifier FT CNTL Plant with 81% Coal / 19% NG Feed and CO_2 Capture (Parallel Indirect Reforming)
- Case 3FT - GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture (Series Indirect Reforming)

Figure 4-1
Reference Case: Simplified BFD - Shell SCGP Gasifier FT CNTL Plant with 100% Illinois No. 6 Coal Feed and CO_2 Capture

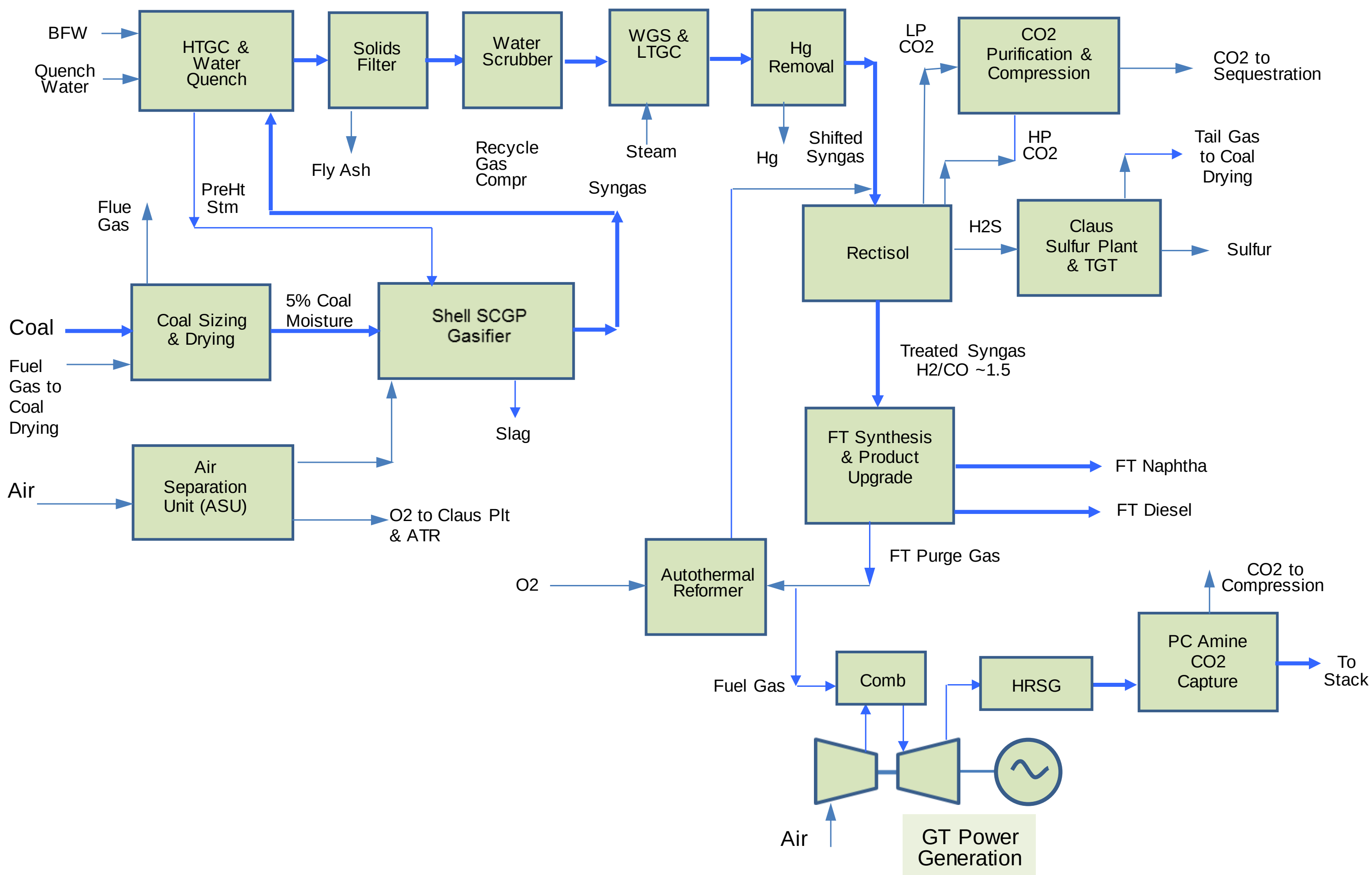


Figure 4-2
Case 1FT: Simplified BFD - GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture

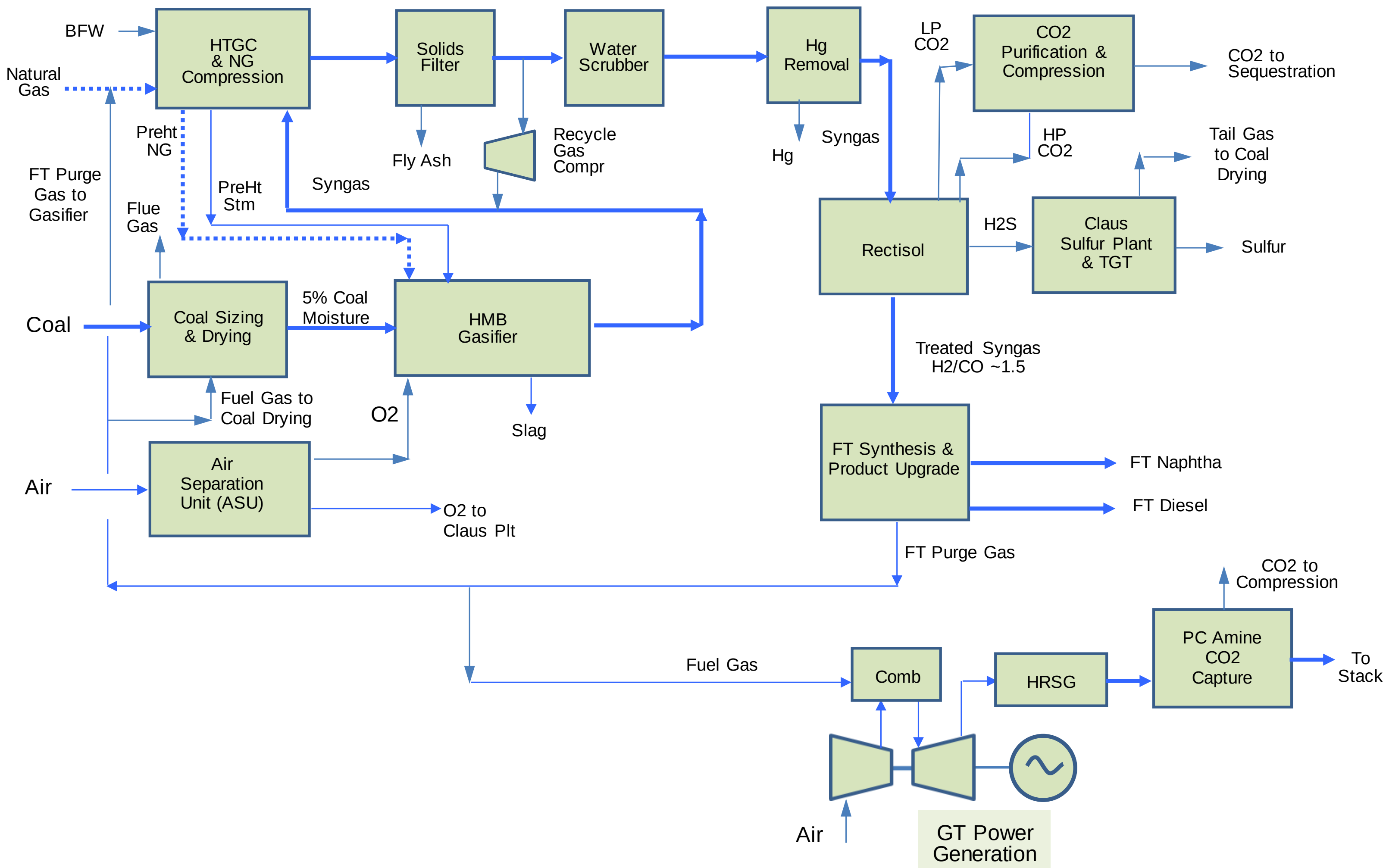


Figure 4-3
Case 2FT: Simplified BFD - GTI HMB Gasifier FT CNTL Plant with 81% Coal / 19% NG Feed and CO_2 Capture (Parallel Indirect Reforming)

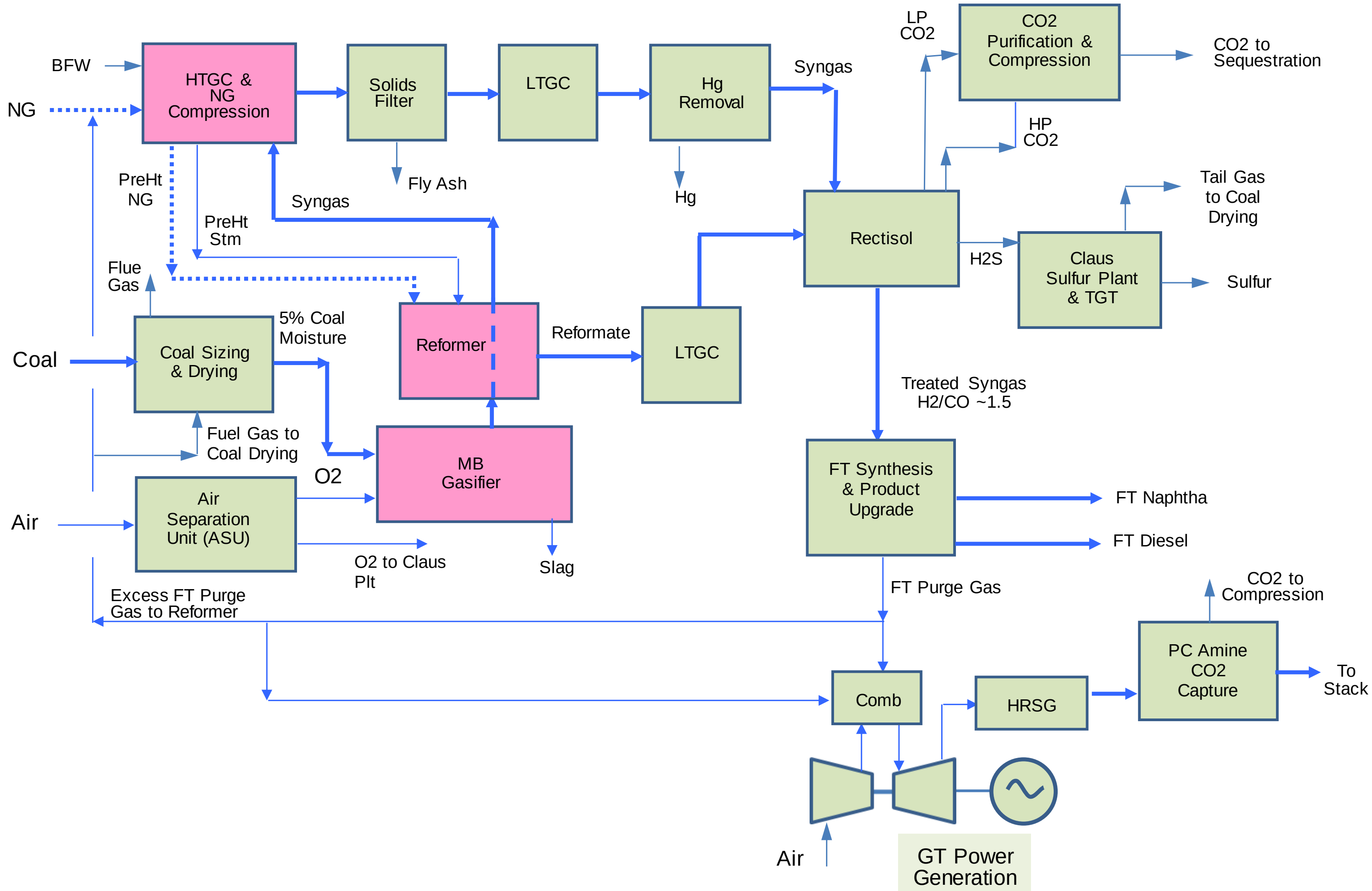
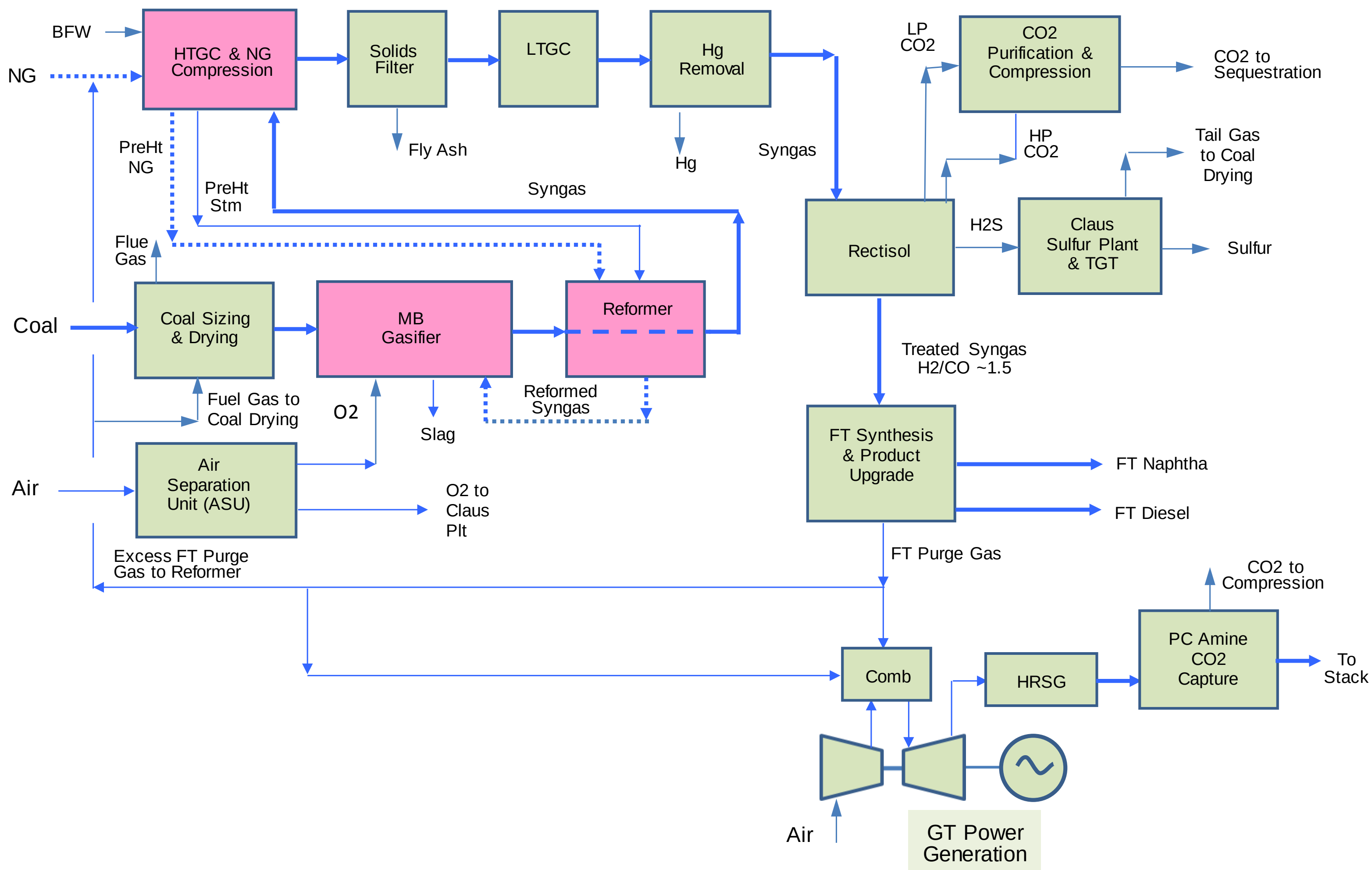


Figure 4-4
Case 3FT: Simplified BFD - GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture (Series Indirect Reforming)



4.1.1. Reference FT CNTL Plant with CO_2 Capture

The reference FT CNTL plant for the techno-economic analysis is based on the Shell gasifier with 100% Illinois No. 6 coal feed and with CO_2 capture. It is designed to produce a nominal 50,000 BPD of FT diesel and naphtha. A simplified Block Flow Diagram (BFD) for the Reference plant is shown in Figure 4-1.

Both the Shell gasification technology (SCGP) for syngas production and the Fischer-Tropsch synthesis process for liquid fuels production have been proven on commercial scale and are considered technologically matured. Hence, their overall performance and costs can be estimated at a high confidence level.

The reference plant gasification section consists of the following units:

- Coal Handling
- Coal Prep, Drying & Feed
- Air Separation Unit (ASU)
- Shell SCGP Gasifier System
- High Temperature Gas Cooling and Steam Generation
- Gas Cleaning (Syngas Scrubbing , Particulate Filters, and Hg Removal)
- Water Gas Shift (WGS)
- Rectisol AGR for H_2S and CO_2 Removal
- CO_2 Compression and Purification Facilities
- Sour Water Stripping
- Sulfur Recovery and Tail Gas Treating

For the reference FT synthesis section, Nexant proposed to use its own iron-catalyst slurry reactor FT design for the GTI study. The current DOE Report 1477 did not separately list the LP fuel gas generations, as well as the FT internal upgrading furnaces LP fuel consumptions. In addition, we were not able to identify FT internal CW loads. In order to facilitate our work on the GTI FT TEA, we proposed to use our own FT design which was based on past projects data from commercial licensors. A comparison between the DOE and the Nexant 50,000 BPSD design are listed in the following table.

Table 4-1
Comparison of DOE FT with Past Nexant's FT Performance and Cost Estimates

	5/12/2014 DOE Report 1477	Past Nexant Estimate
Coal Feed to Shell Gasification:		
AR Illinois #6 Coal Feed, STPD	21,006	23,469
% Moisture in AR Coal	11.1	13.5
AR Coal Fd Carbon Content, STPD	13,391	14,495
% Carbon in Coal	63.8	61.8
FT Liq Production:		
Diesel Product, BPSD	35,230	36,779
C5+ Naphtha Product, BPSD	14,762	12,620
Total C5+ FT Product, BPSD	49,992	49,398
Del/Nap Vol Split	70.5%/29.5%	74.5%/25.5%
Diesel Product, lb/hr	391,830	424,180
C5+ Naphtha Product, lb/hr	147,680	127,090
Total C5+ FT Product, lb/hr	539,510	551,270
Power Production:		
Generation, MWe	427	790
Internal Consumption, MWe	423	648
Net Export, MWe	5	142
Captured CO2 Product:		
STPD CO2	26,405	N/A
STPD Carbon in CO2	7,206	N/A
% Coal Carbon Captured	53.8	N/A
FT Block Information:		
No. of Parallel FT Synthesis Trains	4	4
No. of Reaction Stages per FT Train	2	1
No. of Reactors per FT Train	3	1
Total No. of FT Reactors	12	4
FT B/L Pressure, psia	360	384
FT B/L Temp, deg F	600	77
FT Rx Inlet Temp, deg F	343	~250
FT Rx Outlet Temp, deg F	487	~450
B/L Feed H2/CO Molar Ratio	0.73	1.5
FT Rx Inlet H2/CO Molar Ratio	1.02	1.4
B/L Feed CO2/CO Molar Ratio	0.017	0.08
FT Rx Inlet CO2/CO Molar Ratio	0.017	0.40
B/L Feed H2O/CO Molar Ratio	0.000	0.000
FT Rx Inlet H2O/CO Molar Ratio	0.107	0.006
FT Block CapEx, 2011 $MM:		
FT Synthesis & H2O Processing	340	472
FT Prod Upgrad (HTU & HCU)	60	252
Amine CO2 Abs & Regen	90	Not Incl
FT H2 Recovery	48	43
FT Autothermal Reformer	0	0
FT HP Fuel Gas Compression	28	47
Amine CO2 Abs & Regen	4	Not Incl
FT Block Total Bare Erect Cost	570	814
FT Block E/CM/HO/Fee Cost	55	81
FT Block Process Contingency	146	122
FT Block Project Contingency	191	0
FT Block Total Plant Cost	963	1,018

While the Nexant design shown did not include CO_2 recovery from FT, it will be modified to recover post combustion (PC) CO_2 from the HRSG exhaust. Also, overall CTL plant power production, consumption and export will be modified to be consistent with the DOE report 1477 reference design. We believe Nexant's traditional iron-based FT design with feed H_2/CO of 1.5 will also highlight GTI's mix feed gasification technology better than for feed H_2/CO ratio at less than 1.0.

The iron based FT synthesis process requires a feed gas H_2/CO ratio of 1.5. The syngas exiting the Shell SCGP gasifier with 100% coal feed typically has a H_2/CO ratio of approximately 0.45 which is below the required 1.5 ratio for the FT synthesis reactors. After the gasifier syngas is cooled, filtered and scrubbed, it undergoes catalytic CO shift conversion in the presence of steam (WGS) to increase the H_2/CO ratio of the syngas to 1.5 for FT synthesis.

The shifted syngas from the WGS reactors is cooled and then sent to mercury removal before it is processed in the Rectisol acid gas removal unit to remove H_2S and CO_2 from the syngas. The syngas is then fed to the Fischer-Tropsch synthesis section with essentially all of the sulfur and contaminants removed from the syngas.

Fischer-Tropsch synthesis process has been commercially demonstrated (e.g. SASOL) and can operate with either iron catalyst or the more advance cobalt catalyst. Compared to the iron catalyst, cobalt catalyst is more active but less tolerant to contaminants. The iron catalyst has been in operation with coal-based syngas feed since 1993 while the cobalt catalyst operating experiences are primarily with cleaner natural gas-based syngas feeds. For the HMB techno-economic study, iron catalyst is chosen because of its proven performance for coal derived syngas. The FT synthesis products are typically upgraded by naphtha and diesel hydrotreating , and wax hydrocracking technology to maximize liquid fuel yields and improve the fuel properties of the FT liquid fuels. These upgrading processes are proven technology widely used in the petroleum refining industries.

The Reference synthesis and product upgrade sections are consisted of the following major blocks:

- Fischer-Tropsch synthesis reactor
- Primary product recovery(fractionation)
- Hydrotreating and Hydrocracking
- Heavy Ends Recovery (HER)
- Autothermal Reformer (ATR)

The steam and power generation sections are consisted of the following blocks:

- Steam system
- Combustion Turbine Power Generation (GTG)
- HRSG, Post Combustion (PC) Amine CO_2 Capture, Ducting and Stack
- Steam Turbine Power Generation (STG)

The balance of plant to support the gasification and FT synthesis / products upgrading systems consists of the following major blocks:

- Cooling Water Systems
- BFW/Condensate System
- Slag Recovery and Solids Handling
- Electrical Distribution
- Waste Water Treating

The FT CNTL plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 90 percent.

4.1.2. Case 1FT: GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture

Case 1FT is designed to take advantage of the dual feed capability of the GTI HMB gasifier design by using an optimum coal/NG feed mix to generate in the HMB gasifier the required H_2/CO ratio of 1.5 for the iron based FT synthesis. This will eliminate the need for water gas shift reactors and hence reduce the cost of the FT CNTL plant. The FT CNTL plant is designed to produce a nominal 50,000 BPD of FT diesel and naphtha. A simplified block flow diagram is shown in Figure 4-2.

Case 1FT is the utilization of the dual feed capability of the GTI HMB gasifier design in its simplest form, where coal and natural gas/steam mixture are fed directly to the HMB gasifier to produce FT syngas. FT tail gas is recycled back to the HMB gasifier for conversion. No external reformer is used. It is a variation of the direct reforming concept where gasification and steam reforming take place simultaneously in the HMB gasifier. Sufficient coal is combusted to provide for the reforming duty and to bring the final syngas exit temperature to 2,600°F. The high exit temperature reduces the high methane slippage that normally occurs in direct reforming with lower reforming temperatures.

The Case 1FT gasification section consists of the following units:

- Coal Handling
- Coal Prep, Drying & Feed
- Air Separation Unit (ASU)
- GTI HMB Gasifier System
- High Temperature Gas Cooling and Steam Generation
- Gas Cleaning (Syngas Scrubbing , Particulate Filters, and Hg Removal)
- Rectisol AGR for H_2S and CO_2 Removal
- Post Combustion (PC) Amine CO_2 Capture
- CO_2 Compression and Purification Facilities
- Sour Water Stripping
- Sulfur Recovery and Tail Gas Treating

The iron based FT synthesis process configuration is similar to the reference Shell case except the autothermal reformer (ATR) is not used to convert the hydrogen, CO, and hydrocarbons in the FT tail gas to syngas. The FT tail gas is recycled directly to the HMB gasifier.

The Case 1FT FT synthesis and product upgrade sections are consisted of the following major blocks:

- Fischer-Tropsch synthesis reactor
- Product recovery and upgrade
- Heavy Ends Recovery (HER)
- Reaction Water Treatment

The balance of plant to support the gasification and FT synthesis / products upgrading systems consists of the same units as the Reference case including the Post Combustion (PC) Amine CO_2 Capture unit.

The Case 1FT FT CNTL plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 90 percent.

4.1.3. Case 2FT: GTI HMB Gasifier FT CNTL Plant with 81% Coal / 19% NG Feed and CO_2 Capture (Parallel Indirect Reforming)

Case 2FT is designed to generate the required H_2/CO ratio of 1.5 for the iron based FT synthesis using the parallel indirect reforming configuration. A simplified block flow diagram is shown in Figure 4-3.

This configuration utilizes an external steam methane catalytic reformer where natural gas and/or FT tail gas is reformed with steam. However, instead of returning the reformer syngas to the HMB gasifier, the relatively clean reformer syngas is cooled and processed for contaminant removal separately from the gasifier syngas. The reformer syngas contains very low concentration of particulates, mercury and sulfur species and is sent to the CO_2 removal section only in the Rectisol AGR. The gasifier syngas is cooled and cleaned to remove contaminants such as particulates, mercury, chlorides, ammonia and various sulfur species to minimize contaminants in the FT feed. The external steam methane reformer duty is provided by the 2,600°F syngas exiting the GTI HMB gasifier. The syngas exiting the reformer is at 1,500°F has a H_2/CO ratio of ~3. The HMB gasifier using an 81% coal / 19% NG co-feed and a reformer feed steam/C ratio of 1.3 mole/mole is capable of producing the required FT feed H_2/CO ratio of 1.5. The FT feed is a mix of the gasifier and reformer syngas.

As in Case 1FT, by generating the required FT feed syngas H_2/CO ratio of 1.5 in the HMB gasifier, Case 2FT eliminates the need for water gas shift reactors and reduces the cost of the FT CNTL plant. The FT CNTL plant is designed to produce a nominal 50,000 BPD of FT diesel and naphtha.

The Case 2FT gasification section consists of the following units:

- Coal Handling
- Coal Prep, Drying & Feed
- Air Separation Unit (ASU)
- GTI HMB Gasifier System
- Steam Methane Reformer
- High Temperature Gas Cooling and Steam Generation
- Gas Cleaning (Syngas Scrubbing , Particulate Filters, and Hg Removal)
- Rectisol AGR for H_2S and CO_2 Removal
- CO_2 Compression and Purification Facilities
- Sour Water Stripping
- Sulfur Recovery and Tail Gas Treating

The iron based FT synthesis process configuration is similar to the reference Shell case except the steam methane reformer is used instead of the autothermal reformer (ATR) to convert the hydrogen, CO, and hydrocarbons in the FT tail gas to syngas. The FT tail gas is recycled to the steam methane reformer where it replaces and reduces the natural gas consumption in the steam methane reformer.

The Case 2FT FT synthesis and product upgrade sections are consisted of the following major blocks:

- Fischer-Tropsch synthesis reactor
- Primary product recovery (fractionation)
- Hydrotreating and Hydrocracking
- Heavy Ends Recovery (HER)
- Steam system
- Combustion Turbine Power Generation (GTG)
- HRSG, Post Combustion (PC) Amine CO_2 Capture, Ducting and Stack
- Steam Turbine Power Generation (STG)

The balance of plant to support the gasification and FT synthesis / products upgrading systems consists of the same units as the Reference case.

The Case 2FT FT CNTL plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 90 percent.

4.1.4. Case 3FT: GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed, Reformer and CO_2 Capture (Series Indirect Reforming)

Case 3FT is designed to generate the required H_2/CO ratio of 1.5 for the iron based FT synthesis using the series indirect reforming configuration. A simplified block flow diagram is shown in Figure 4-4.

This configuration utilizes an external steam methane catalytic reformer where natural gas and/or FT tail gas is reformed with steam. The reformer duty is provided by the 2,600°F syngas exiting the GTI HMB gasifier. Case 3FT differs from Case 2FT in that the reformer syngas is returned to the HMB gasifier through the dual feed gasifier burners carrying with it the recuperated heat from the gasifier syngas. The heat recuperation improves plant efficiency and is possible due to the unique feature of the GTI dual feed gasifier for handling gaseous feed. The syngas exiting the reformer is at 1,500°F has a H_2/CO ratio of ~3. The HMB gasifier using a 55% coal / 45% NG co-feed and a reformer feed steam/C ratio of 1.4 mole/mole is capable of producing the required FT feed H_2/CO ratio of 1.5.

The gasifier syngas is cooled and cleaned to remove contaminants such as particulates, mercury, chlorides, ammonia and various sulfur species to minimize contaminants in the FT feed.

As in Case 1FT and 2FT, by generating the required FT feed syngas H_2/CO ratio of 1.5 in the HMB gasifier, Case 3FT eliminates the need for water gas shift reactors and reduces the cost of the FT CNTL plant. The FT CNTL plant is designed to produce a nominal 50,000 BPD of FT diesel and naphtha.

The Case 3FT gasification section consists of the following units:

- Coal Handling
- Coal Prep, Drying & Feed
- Air Separation Unit (ASU)
- GTI HMB Gasifier System
- Steam Methane Reformer
- High Temperature Gas Cooling and Steam Generation
- Gas Cleaning (Syngas Scrubbing , Particulate Filters, and Hg Removal)
- Rectisol AGR for H_2S and CO_2 Removal
- CO_2 Compression and Purification Facilities
- Sour Water Stripping
- Sulfur Recovery and Tail Gas Treating

The iron based FT synthesis process configuration is similar to the reference Shell case except the steam methane reformer is used instead of the autothermal reformer (ATR) to convert the hydrogen, CO, and hydrocarbons in the FT tail gas to syngas. The FT tail gas is recycled to the steam methane reformer where it replaces and reduces the natural gas consumption in the steam methane reformer.

The Case 3FT FT synthesis and product upgrade sections are consisted of the following major blocks:

- Fischer-Tropsch synthesis reactor
- Primary product recovery (fractionation)
- Hydrotreating and Hydrocracking
- Heavy Ends Recovery (HER)
- Steam system
- Combustion Turbine Power Generation (GTG)
- HRSG, Post Combustion (PC) Amine CO_2 Capture , Ducting and Stack
- Steam Turbine Power Generation (STG)

The balance of plant to support the gasification and FT synthesis / products upgrading systems consists of the same units as the Reference case.

The Case 3FT FT CNTL plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 90 percent.

5. REFERENCE CASE: SHELL FT CNTL PLANT WITH 100% COAL FEED AND CO2 CAPTURE

5.1.Process Overview

The reference FT CNTL plant for the techno-economic analysis is based on the Shell gasifier with 100% Illinois No. 6 coal feed and with CO_2 capture. It is designed to produce a nominal 50,000 BPD of FT diesel and naphtha. A simplified Block Flow Diagram (BFD) for the Reference plant is shown in Figure 4-1.

The iron based FT synthesis process requires a feed gas H_2/CO ratio of 1.5. The syngas exiting the Shell SCGP gasifier with 100% coal feed typically has a H_2/CO ratio of approximately 0.45 which is below the required 1.5 ratio for the FT synthesis reactors. After the gasifier syngas is cooled, filtered and scrubbed, it undergoes catalytic CO shift conversion in the presence of steam (WGS) to increase the H_2/CO ratio of the syngas to 1.5 for FT synthesis.

The shifted syngas from the WGS reactors is cooled and then sent to mercury removal before it is processed in the Rectisol acid gas removal unit to remove H_2S and CO_2 from the syngas. The syngas is then fed to the Fischer-Tropsch synthesis section with essentially all of the sulfur and contaminants removed from the syngas.

For the HMB techno-economic study, iron catalyst is chosen because of its proven performance for coal derived syngas. The FT synthesis products are typically upgraded by naphtha and diesel hydrotreating, and wax hydrocracking technology to maximize liquid fuel yields and improve the fuel properties of the FT liquid fuels.

5.2. Process Description

5.2.1. Coal Sizing and Handling

The Illinois No. 6 coal is delivered to the site by 100-ton rail cars. It is unloaded into 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.

5.2.2. Coal Drying

The Illinois No. 6 coal contains 11.12 wt% total moisture on an as-received basis. The coal is crushed and dried in the coal mill and delivered to a surge hopper. The heat for drying is provided by the combustion of tail gas from the FT unit. It is assumed that the coal is dried to 5% moisture for smooth feed through the pneumatic conveying system.

5.2.3. ASU

Air separation consists of a multiple train cryogenic separation unit, supplying 95% purity high-pressure oxygen to the Shell gasifier and the Claus plant. It also provides 99.9% purity nitrogen for the Rectisol unit and for other plant services. Nitrogen is also needed for instrument air back-up and shut down operations such as purging of the gasifier.

The battery limit conditions for the ASU products are summarized below:

Table 5-1
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	164	64

A total of 21,808 tons/day of 95mol% oxygen is required for the reference case. The Shell gasifier requires 19,710 tons/day with the balance going to the ATR.

5.2.4. Gasification

Two trains of five Shell dry feed gasifiers each are used to process a total of 23,000 tons/day of as-received Illinois No. 6 coal. These gasifiers operate at 545 psia. Coal is gasified with 95% oxygen from the ASU to produce syngas containing H_2, CO, CO_2, H_2O, NO_x, SO_x and other products of coal gasification. The gasifier membrane wall is cooled by steam generation to create a protective ash layer over the membrane to maintain the gasification temperature at ~2,600°F. High carbon conversion (above 99%) is obtained in the gasifier, and the high temperature ensures that essentially no organic components heavier than methane are in the raw syngas. The insulation provided by the slag layer in the gasifier minimizes heat losses.

5.2.5. 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 lock hopper for dewatering. The water is clarified and reused as makeup to the water scrubber. The dried slag is stored for disposal.

5.2.6. Syngas Cooling & Particulate Filters

The raw syngas from the gasifier is quenched to below the ash melting point (~2,298°F) by cooled recycled syngas. It is then water quenched to 685°F before entering the ceramic particulate filters and cyclones. Any remaining particulate matters in the syngas will be removed by these filters and cyclones. The filtered syngas is then cooled to 450°F through a series of steam generators generating steam for the steam cycle. The cooled syngas enters a water scrubber to remove any chlorides and remaining particulates. The scrubbed syngas is sent to WGS. Part of the scrubber bottoms is used for slag water bath makeup.

Make-up water is continuously added to the wet particulate removal unit to control the concentration of contaminants in the blowdown stream. The contaminated water is sent to the sour water stripper to recover the contaminants.

5.2.7. Water Gas Shift

In order to achieve the required H_2/CO ratio of 1.5 for FT feed, the raw syngas must undergo a catalytic CO shift conversion (WGS) with steam. Approximately 10% of the syngas is bypassed around the WGS to control the WGS conversion. Two reactor stages in series are used for the specified CO_2 capture requirements. Inter-stage cooling by steam generation is required to control the exothermic temperature rise in the reactors.

The WGS reactors are located downstream of the water scrubber and ahead of the AGR. They contain sulfur tolerant shift catalysts for the WGS and COS hydrolysis reactions. As the syngas still contains H_2S and COS, the concept is called “sour shift” in general. The shift conversion of CO is based on the exothermic reaction

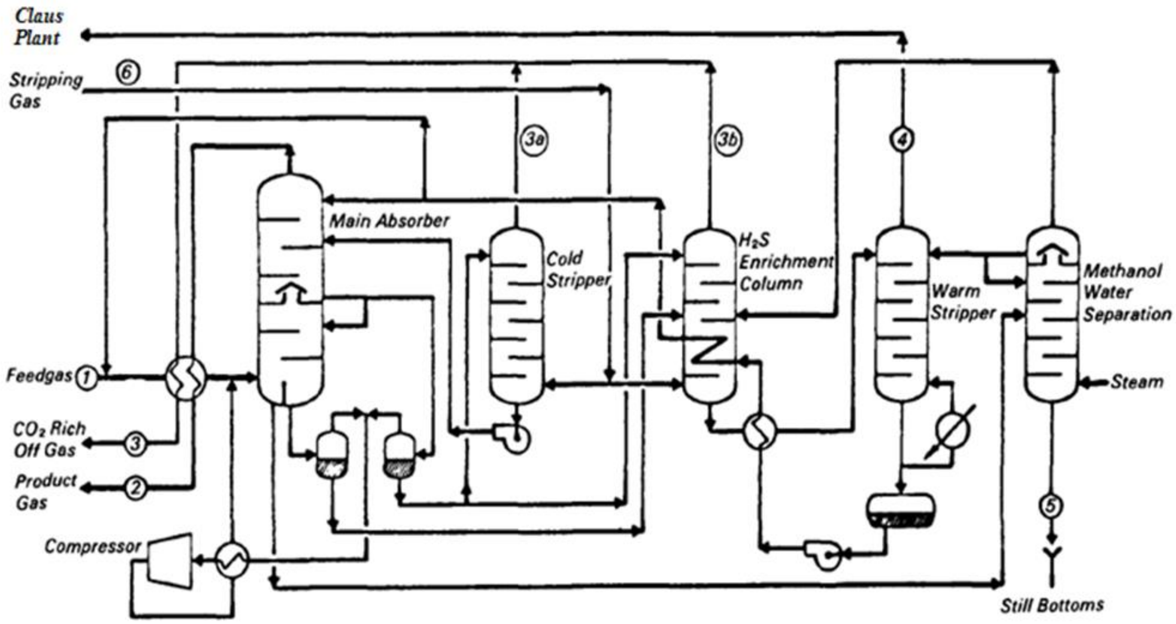
$$CO + H2O \leftrightarrow H2 + CO2 \quad \text{at approximately } 500°F$$

5.2.8. Mercury Removal

The shifted syngas from the WGS is cooled to 100°F and sent to a packed carbon bed vessel designed to treat the cooled syngas for mercury removal. The beds are designed for superficial gas velocity of 1 ft/sec and design residence time of 20 seconds.

5.2.9. Rectisol Acid Gas Removal (AGR)

The Rectisol AGR is designed for 90% CO_2 recovery and a product CO_2 purity of 95%. An acid gas stream is also produced with a minimum H_2S content of 25%. The CO_2 product is sent to purification and is compressed for sequestration. The acid gas stream is processed in a Claus/TGTU to control sulfur emission by recovering sulfur as elemental sulfur. The treated syngas is sent to FT synthesis unit. A typical schematic of the Rectisol unit is shown below.



Source: EPA

The Rectisol AGR unit receives syngas from mercury removal at about 100°F. Rectisol AGR utilizes a refrigerated methanol solvent which has high capacity for H_2S and CO_2 and is also capable of removing COS in the syngas. This added capability eliminates the need for the COS hydrolysis pretreatment step.

Before it enters the main Rectisol absorber, the syngas is chilled by heat exchange with the Rectisol products. The H_2S, COS and CO_2 components of the syngas are absorbed by the chilled lean methanol solvent in the bottom section of the absorber. The sulfur free syngas is then washed in the upper stages of the absorber to remove the remaining CO_2. The H_2S/CO_2 rich solvent from the absorber bottom is flashed in a medium pressure flash drum to recover H_2 and CO containing fuel gas. The fuel gas which is also rich in CO_2 is compressed and recycled back to the absorber or purged to the fuel system. The MP flash drum bottoms liquid is sent to the medium pressure stripper where CO_2 is stripped from the rich methanol solvent. The bottoms liquid from the MP stripper and a side cut taken from the bottom of the main absorber are sent to a low pressure (LP) stripper where additional CO_2 are removed from the methanol solvent by stripping with nitrogen from the ASU. The H_2S rich methanol solvent stream from the LP stripper is sent to a reboiled stripper where H_2S rich acid gas is produced and is sent to the Claus sulfur plant for sulfur recovery. The CO_2 rich gas from the MP and LP flashes are sent to CO_2 purification and compression.

Additional CO_2 removal is required to achieve the < 10% CO_2 emissions target. A PC amine CO_2 removal unit is provided to remove > 85% of the CO_2 from the HRSG effluent gas.

5.2.10. Claus/TGTU Plant

Since oxygen is available from the ASU, oxygen-blown Claus sulfur plant is selected for recovering sulfur from the process acid gas streams. The process streams include H_2S rich streams from the AGR, TGTU tail gas recycle and the Sour Water Stripper off-gas. The oxygen-blown Claus process can provide operating flexibility and lower cost than conventional Claus process. The H_2S rich feed streams are first combusted in the Claus sulfur reaction furnace to form SO_2 which is then converted to elemental sulfur in the catalytic reactors. The Claus plant tail gas is further treated in a tail gas treating unit (TGTU) to meet the SO_2 emission target. Three Claus catalytic stages and a tail gas treating unit (TGTU) are required to achieve the sulfur recovery efficiency of 99.8%.

5.2.11. PSA Hydrogen Purification

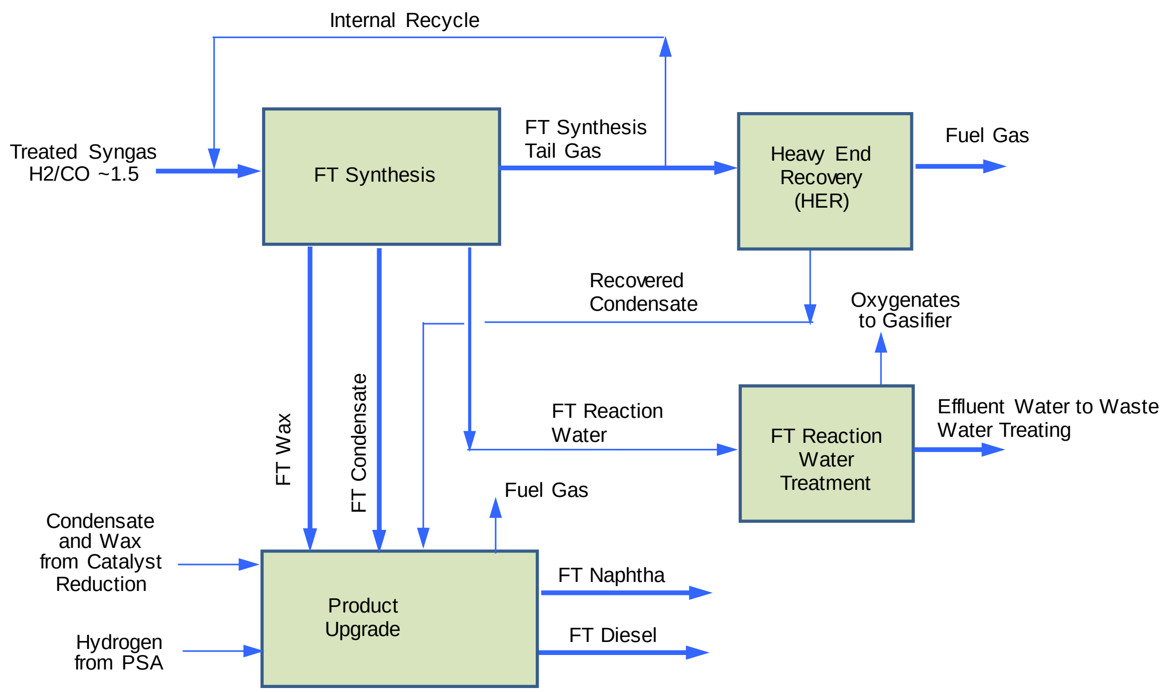
A slipstream of treated syngas from the Rectisol Acid Gas Removal Unit is purified in the PSA unit to recover 99.99% hydrogen. The PSA unit is designed for 37 MMSCFD of purified hydrogen. The product hydrogen is mainly used in the naphtha hydrotreater and the wax hydrocracker units for the hydro-processing of the Fischer-Tropsch reactor liquid. A portion of the hydrogen is also used periodically for catalyst reduction in the FT unit. The residual gas from the PSA unit is sent to the LP fuel system

5.2.12. Fischer-Tropsch Synthesis

The FT synthesis unit is designed to operate at 413 psia and converts 213,276 lbmol/h of treated syngas into 38,053 BPD of FT diesel and 13,057 BPD of FT naphtha for the Shell Reference case. Part of the tail gas generated by the FT synthesis and upgrading units are recycled to improve conversion and the remaining fuel gas is sent to the gas turbine for power generation. Only sufficient power is generated for the FT CNTL plant with net power export minimized.

The Fischer-Tropsch Synthesis Unit converts treated syngas from the Rectisol AGR to raw synthesis products of wax, hydrocarbon condensate, tail gas, and reaction water. The wax and hydrocarbon condensate streams are upgraded in the product upgrading units into final FT naphtha and diesel products. The tail gas from the FT synthesis reactors is sent to a Heavy End Recovery (HER) unit to recover mainly pentane and heavier components. The tail gas stream from the HER Unit is recovered and used as fuel gas. The reaction water by-product stream is treated in the water treatment unit. Figure 5-1 is a simplified BFD of the FT synthesis and product upgrading plant.

Figure 5-1
BFD – FT Synthesis and Product Upgrading



The Fischer-Tropsch synthesis catalyst is activated in a catalyst reduction unit. The catalyst reduction process is a batch operation requiring pure hydrogen and also syngas to reduce and condition the catalyst. During the reduction and conditioning steps, FT wax and condensate are produced, as well as reaction water. The products are sent to their respective processing units (product upgrading and FT reaction water treatment).

The Fischer-Tropsch plant is consisted of the following major units:

- Fischer-Tropsch synthesis reactors
- Product recovery and upgrade
- Heavy Ends Recovery (HER)
- Reaction heat recovery and steam generation system

Fischer-Tropsch Synthesis Reactors

Treated syngas from the gasifier Rectisol AGR plant is combined with recycle gas from the recycle compressor and preheated before entering the FT reactors. The FT reactors operate at about 410 psia. The feed gas, comprising mostly of CO and H_2, is converted to raw synthesis products of wax, hydrocarbon condensate and tail gas via the Fischer-Tropsch synthesis reaction. Reaction water byproducts are also produced.

The wax product from the reactor system is cooled and filtered through a wax separation unit.

The reactor overhead stream is cooled and separated into three phases

- Hydrocarbon condensate
- Reaction water
- Vapor unreacted feed and vapor products

Hydrocarbon condensate is sent to the product upgrading unit for recovery and upgrading. The reaction water stream is sent to the reaction water treatment unit

Part of the overhead vapor from the separator is used as recycle to increase conversion in the FT reactor, while the balance is fed to the HER to recover the pentane and heavier condensates. The tail gas is sent to the fuel gas header and is used to drive gas turbines in the power plant.

The major products of the unit are:

- Wax to product upgrading
- Un-stabilized hydrocarbon condensate to product upgrading
- Reaction water to reaction water treatment unit
- Tail gas to fuel gas system

Reaction Heat Recovery & Steam System

The process steam system removes the net heat of reaction from the FT reactor by circulating boiler feed water (BFW) through internal coils in the reactor where it is partially vaporized to generate steam.

Product Upgrading Unit

The product work-up unit consists of the following operations:

- Condensate hydrotreating
- Wax hydrocracking
- Product fractionation.

The unit converts the hydrocarbon condensate and wax streams from the FT synthesis unit into FT naphtha and a diesel blend. The operation also produces light end hydrocarbon materials, which are consumed as fuel within the plant.

CO_2 Stripping

The un-stabilized condensate from the FT synthesis unit and recovered condensate from the HER Unit is sent to a CO_2 stripper column. In the column, CO_2 is stripped from the liquid feed and the feed is stabilized.

Condensate Hydrotreating

The stabilized hydrocarbon condensate is hydrotreated to saturate the olefins and remove the oxygenate-containing compounds. Product from the hydrotreater reactor is combined with the hydrocracker product for fractionation.

The recycle gas circuit supplies both the condensate and wax reactor loops with hydrogen-rich gas for hydrotreating and hydrocracking. Hydrogen quench is used between the catalyst beds of the reactors to control the catalyst bed temperature.

Wax Hydrocracking

Wax from the FT synthesis unit is hydrocracked in the wax hydrocracker reactors to yield naphtha and diesel as final products.

Product Fractionation

The product fractionator separates the hydrotreater and hydrocracker reactor products by steam stripping. The following intermediate products streams are produced.

- Off-gas to fuel gas system
- Un-stabilized naphtha
- Diesel
- Unreacted bottoms recycle
- Waste water

Naphtha Stripping

Un-stabilized naphtha is separated into hydrocarbon light ends and naphtha in the naphtha stabilizer column. The light ends are fed to the fuel system and the product naphtha is pumped to storage.

Reaction Water Treatment Unit

The reaction water from the FT synthesis unit contains oxygenates, including alcohols, ketones, aldehydes and carboxylic acids, which are by-products of the synthesis reaction. The effluent water treatment unit removes the non-acid chemicals so that the effluent water can be sent to bio-treatment and disposal. The chemicals are recycled to gasification for disposal and their energy content.

5.2.13. Autothermal Reforming (ATR)

The FT tail gas contains light hydrocarbons that can be converted to H_2 and CO and recycled to the FT synthesis section as feed. The ATR uses oxygen to partially oxidize the light hydrocarbons and catalytically reforms these light hydrocarbons into H_2 and CO with steam. Typical reforming temperature for the ATR is about 1,800°F.

5.2.14. CO_2 Compression and Dehydration

MP and LP flashed CO_2 from the Rectisol AGR plant are at the battery limit conditions of 42 and 19 psia respectively and at 80°F. Along with the CO_2 from the PC amine unit, they are purified to sequestration CO_2 specification and compressed to supercritical condition of 2,215 psia using a multi-stage, intercooled compressor. The CO_2 stream is dehydrated to a dew point of -40°F at the appropriate inter-stage pressure using a thermal swing adsorptive dryer.

5.2.15. Gas Turbine

The gas turbine generator selected is a GE SG6FA class turbine with a nominal ISO gross GT output of 95 MW. Nitrogen from the ASU is used for dilution to limit NO_x formation and to adjust the syngas LHV to 115-132 Btu/Scf. Inlet air is compressed to a pressure ratio of 15:1 for the GT combustion process. Hot combustion products are expanded in the gas turbine expander with an exhaust temperature of around 1,028°F. Three GTs are used for the reference case for a total of 260 MW GT output.

5.2.16. Steam Turbine and HRSG

The 1,028°F GT exhaust is cooled in the HRSG by generating HP, IP and LP steams for the steam turbines (ST) and process users. The cooled GT flue gas exits the HRSG at 191°F and is vented to the atmosphere through a stack. HP steam is used in the ST for power generation. LP exhaust steam from the last ST stage is condensed. The condensers operate at 0.698 psia with a corresponding condensing temperature of 90°F.

The condensates are collected and send 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 steam generators.

5.2.17. BOP

Raw Water System

Raw water system supplies cooling tower makeup, demineralizer water makeup, fire protection system water and potable water requirements. The water source is 50 percent from potable water and 50 percent from groundwater.

The demineralizer makeup system consists of two 100 percent trains, each with a 100% capacity activated carbon filter, primary cation and anion exchanger, mixed bed exchanger, recycle pump, and regeneration equipment. It provides demineralized water for HRSG BFW makeup.

The fire protection system provides pressurized water to the fire hydrants, hose stations and fire suppression sprinkler systems.

Accessory Electric Plant

The accessory electric plant is consisted 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.

Instrumentation and Control

A plant wide distributed control system (DCS) is provided.

5.3.Shell Reference Case Performance

The Shell reference case performance is shown in Table 5-2. The power generation and the auxiliary loads were calculated based on Nexant's modeling of the reference case. The FT CNTL design is based on minimizing the net power export. Sufficient FT tail gas is used in the gas turbines to generate power to satisfy the auxiliary loads. The balance of the FT tail gas is converted in the ATR to generate additional H_2/CO for the FT feed.

Table 5-2
Power Generation and Auxiliary Load Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Reference Shell Gasifier Case
Gas Turbine Power	259,913
Steam Turbine Power	406,945
TOTAL POWER, kWe	**666,858**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling & Milling	41,712
Slag Handling	2,570
Natural Gas Compressors	0
Gasifier System	11,035
Air Separation Unit Main Air Compressor & Auxiliaries	181,822
Oxygen Compressor	105,306
CO2 Compressor	52,289
Boiler Feedwater Pumps	40,003
Condensate Pump	170
Circulating Water Pump	32,614
Ground Water Pumps	602
Cooling Tower Fans	10,456
Acid Gas Removal	41,768
Claus Plant/TGTU Auxiliaries	2,414
Sour Water Stripper	86
FT Power Requirement	52,208
PSA & TG Recycle	74
ATR w/Feed Pump	13,811
Miscellaneous Balance of Plant	74,919
TOTAL AUXILIARIES, kWe	**663,859**
NET POWER, kWe	**2,998**
CONSUMABLES	
As-Received Coal Feed, lb/hr	1,916,667
Natural Gas Feed, lbs/hr	0
Thermal Input, kWt	6,553,020
Condenser Duty, MMBtu/hr	1,135
Raw Water Withdrawal, gpm	47,729
Raw Water Consumption, gpm	43,501

5.3.1. Carbon Balance

Table 5-3 shows the carbon balance for the Case 1FT GTI HMB gasifier-based FT CNTL plant. Carbon emission is based on the total stack emissions from the process heaters, cogen and SRU/TGTU. CO_2 is recovered from the AGR and the PC amine CO2 recovery units with the balance of carbons going to the FT products. The waste liquid effluent and ash byproducts contain a small amount of carbon.

Table 5-3
Reference Shell Gasifier Case Carbon Balance

OVERALL CARBON BALANCE		
Case 0FT		
Case Description	100% Coal Reference Shell Gasifier Case	
	lbs/hr	**%**
Carbon IN:		
Coal & Flux	1,374,632	99.96
Cogen Combustion Air	601	0.04
Total Carbon In	**1,375,233**	**100.00**
Carbon OUT:		
Process Htr Stack	47,958	3.49
Cogen Stack	25,046	1.82
SRU/TGTU Stack	59,204	4.31
Recovered C from AGR	579,772	42.16
Export FG	0	0.00
AGR Purge H2O	151	0.01
FT Waste H2O	7,868	0.57
Slag & Ash	7,483	0.54
Naphtha + Diesel Products	505,851	36.78
PC Amine CO2 Recovery	141,928	10.32
Convergence Error	-29	0.00
Total Carbon Out	**1,375,233**	**100.00**
Carbon Emission:		
Total Carbon Emissions	140,228	9.61

5.3.2. Water balance

Water makeup and consumptions are shown in Table 5-4. The scrubber water demand is based on a maximum chlorides concentration in the scrubber purge water of 1,000 ppmw.

Table 5-4
Reference Shell Gasifier Case Water Balance

Water Balance (GPM)	**Case 0FT Reference Shell Gasifier Case**				
Water Usage by Area	Water Demand	Internal Recycle	Net Water Demand	Process Water Discharge	Raw Water Withdrawal
(GPM)					
Slag Handling	873		873	-	
SG Scrubber Makeup	13,162		13,162	(12,452)	
SWS	328	(328)	-		
SG Cooling Cond		(325)	(325)		
AGR/SRU/TGU Steam Cond		(4,898)	(4,898)	(195)	
FT Block			-	(1,363)	
BFW	25,177	(6,110)	19,067		
Blowdowns			-	(676)	
Fuel Gas Saturator	980		980		
Cooling Tower	14,594		14,594	(1,911)	
Potable Water + Contingency	48		48	8	
Total Water Usage	55,161	(11,660)	43,501	(16,589)	
Raw Water Treating					
RO/Demin System			(16,073)	(2,832)	18,858
Makeup Cooling Water Treating			(27,428)	(1,444)	28,871
Total Treated Water			(43,501)	(4,276)	47,729
Total	55,161	(11,660)	-	(20,865)	47,729

5.4.Capital Cost

5.4.1. Total Plant Cost

Table 5-5 shows the total plant cost (TPC) summary for the Reference Shell Gasifier Case. Account 4 includes the costs of the gasifier, syngas cooling and ASU.

Table 5-5
Reference Shell Gasifier Case – Total Plant Cost Summary

Total Plant Cost (June 2011)		GTI HMB Case 0FT
Acct. No.	Item/Description	$MM
1	COAL & SORBENT HANDLING	$ 102.74
2	COAL & SORBENT PREP & FEED	$ 548.72
3	FEEDWATER & MISC BOP SYSTEMS	$ 92.24
4	GASIFIER & ACCESSORIES	$ 2,531.10
5	GAS CLEANUP & PIPING	$ 1,172.22
5AA	FT SYNTHESIS AND PRODUCT UPGRADE	$ 1,142.62
5B.2	CO2 Compression & Drying	$ 105.80
6	COMBUSTION TURBINE/ACCESSORIES	$ 147.58
7	HRSG, PC AMINE UNIT, DUCTING & STACK	$ 141.81
8	STEAM TURBINE GENERATOR	$ 97.40
9	COOLING WATER SYSTEM	$ 75.23
10	ASH/SPENT SORBENT HANDLING SYS	$ 112.14
11	ACCESSORY ELECTRIC PLANT	$ 148.15
12	INSTRUMENTATION & CONTROL	$ 37.16
13	IMPROVEMENTS TO SITE	$ 51.10
14	BUILDINGS & STRUCTURES	$ 36.65
		$ 6,542.67

5.5.Operating Costs

Table 5-6 shows the operating costs breakdown for the Reference Shell Gasifier Case.

Table 5-6
Reference Shell Gasifier Case – Operating Cost Breakdown

OPERATING COSTS, 2011 $MM/yr	GTI HMB Case 0FT
FIXED OPERATING COSTS	
Annual Operting Labor Cost	$28.9
Maintainence Labor Cost	$58.8
Administration & Support Labor	$21.9
Property Taxes and Insurance	$130.9
TOTAL FIXED OPERATING COSTS	$240.5
VRIABLE OPERATING COSTS (@90% CF)	
NON-FUEL VARIABLE OPERATING COSTS	
Maintenance Material Cost	$111.0
Water	$18.9
Chemicals	
MU & WT Chemicals	$18.0
Other Chemicals & Catalysts	$18.0
Waste Disposal	$21.6
Power Credits	($1.4)
TOTAL NON_FUEL VARIABLE OPERATING COSTS	$186.1
FUEL (@90% CF)	
Coal	$518.3
Natural Gas	$0.0
TOTAL VARIABLE OPERATING COSTS	$704.4

5.6.Cost of Electricity

Table 5-7 shows a summary of the power output, CAPEX, OPEX, COP and cost of CO_2 capture for the Reference Shell Gasifier Case.

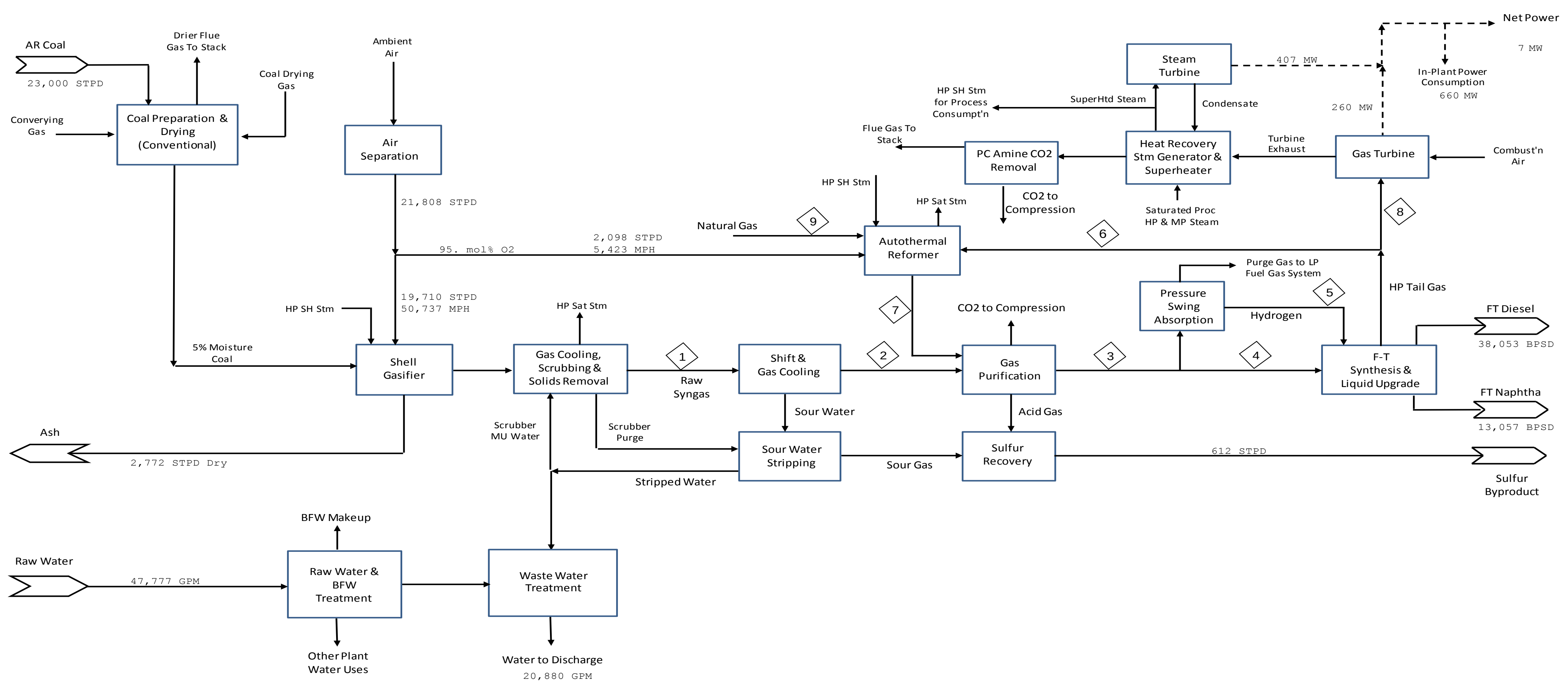
Table 5-7
Reference Shell Gasifier Case – Plant Performance and Economic Summary

	GTI HMB Case 0FT
CAPEX, $MM	
Total Installed Cost (TIC)	$4,625
Total Plant Cost (TPC)	$6,543
Total Overnight Cost (TOC)	$8,078
OPEX, $MM/yr (90% Capacity Factor Basis)	
Fixed Operating Cost (OC_{Fix})	$240
Variable Operating Cost Less Fuel (OC_{VAR})	$186
Fuel Cost (OC_{Fuel})	$518
Power Production, Mwe	
Gas Turbine	259.9
Steam Turbine	406.9
Auxiliary Power Consumption	663.9
Net Power Output	3.0
Power Generated, MWh/yr (MWH)	26,266
COP FT Diesel, excl CO2 TS&M, $/bbl FT diesel	173.6
COP FT Diesel, incl CO2 TS&M, $/bbl FT diesel	185.2
COP FT EPD, excl CO2 TS&M, $/bbl EPD	$168.3
COP FT EPD, incl CO2 TS&M, $/bbl EPD	$179.9
COP FT ECO, excl CO2 TS&M, $/bbl ECO	134.7
COP FT ECO, incl CO2 TS&M, $/bbl ECO	146.2
COP FT Naphtha, excl CO2 TS&M, $/bbl FT Naphtha	120.8
COP FT Naphtha, incl CO2 TS&M, $/bbl FT Naphtha	132.4

5.7.BFD & stream data

A block flow diagram with stream data for the Reference Shell Gasifier Case is shown below.

Figure 5-2
BFD for Reference Shell Gasifier Case



Steam No.	1	2	3	4	5	6	7	8	9
Description	Raw Syngas	Shifted Syngas	Clean Syngas	Syngas to FT	H2 to Upgrade	Tail Gas to Upgrade	Upgraded Gas to AGR	HP Tail Gas to GT	Natural Gas to ATR
Temperature,F	296	100	95	95	95	95	95	95	95
Pressure, psia	505	440	420	413	393	340	440	340	497
Flow, lb-mole/hr									
CO	112,100	68,774	83,875	79,311	0	14,832	15,102	7,122	0
H2	56,121	99,447	122,739	116,735	4,008	20,214	23,292	9,112	0
CO2	7,663	50,989	6,716	6,351	0	11,307	14,921	4,907	0
CH4	49	49	412	389	0	1,598	362	689	0
C2H6	0	0	0	0	0	147	0	63	0
C3H8	0	0	0	0	0	289	0	124	0
C4 +	0	0	0	0	0	327	0	140	0
Ar + N2	3,484	3,484	11,088	10,491	40	7,333	7,604	3,237	0
H2S	1,462	1,462	0	0	0	0	0	0	0
H2O	25,639	484	0	0	0	21	128	9	0
Total, lb-mole/hr	206,650	224,822	224,830	213,276	4,048	56,070	61,409	25,403	0
Total, lb/hr	4,231,089	4,558,491	3,283,712	3,106,559	9,214	1,272,942	1,399,192	573,979	0

Rev No.	Date	Revision	Engineer

Nexant, Inc.
San Francisco, California
DOE/NETL Advanced Gasification Technologies Program
GTI Hybrid Molten Bed Gasifier for Production of High Hydrogen Syngas Project
Overall Block Flow Diagram
Case 0FT (Reference Shell Gasifier)
HMB Gasifier F-T CNTL Plant
100% Coal Feed with CO2 Capture

Job No.	Drawing No.	Revision No.
A02220	BFD - 000	0

6. CASE 1FT: GTI HMB GASIFIER FT CNTL PLANT WITH 55% ILLINOIS NO. 6 COAL / 45% NG FEED AND CO_2 CAPTURE

6.1. Process Overview

Case 1FT is designed to take advantage of the dual feed capability of the GTI HMB gasifier by using an optimum coal/NG feed mix to generate in the HMB gasifier the required H_2/CO ratio of 1.5 for the iron based FT synthesis. This will eliminate the need for water gas shift reactors and hence reduce the cost of the FT CNTL plant. The FT CNTL plant is designed to produce a nominal 50,000 BPD of FT diesel and naphtha. A simplified block flow diagram is shown in Figure 4-2.

Case 1FT is the utilization of the dual feed capability of the GTI HMB gasifier design in its simplest form, where coal and natural gas/steam mixture are fed directly to the HMB gasifier to produce FT syngas. FT tail gas is recycled to the gasifier. No external reformer is used. It is a variation of the direct reforming concept where gasification and steam reforming take place simultaneously in the HMB gasifier. Sufficient coal is combusted to provide for the reforming duty and to bring the final syngas exit temperature to 2,600°F. The high exit temperature reduces the high methane slippage that normally occurs in direct reforming with lower reforming temperatures.

The iron based FT synthesis process configuration is similar to the reference Shell case except the autothermal reformer (ATR) is not used to convert the hydrogen, CO, and hydrocarbons in the FT tail gas to syngas. The FT tail gas is recycled directly to the HMB gasifier.

The steam, power generation section and the balance of plant to support the gasification and FT synthesis / product upgrading systems consists of the same units as the Reference case.

6.2. Process Description

The process descriptions for the various Case 1FT subsystems are identical to those described for the Shell Reference FT CNTL case in Section 5.2, except for the gasification sections dual-fueled gasification section, which runs on a combination of natural gas and coal, as described in Section 6.2.1 below.

6.2.1. ASU

ASU for Case 1FT is similar to the reference case ASU. Case 1FT requires 29,248 tons/day of 95mol% oxygen for the GTI HMB gasifier.

6.2.2. GTI HMB Gasification

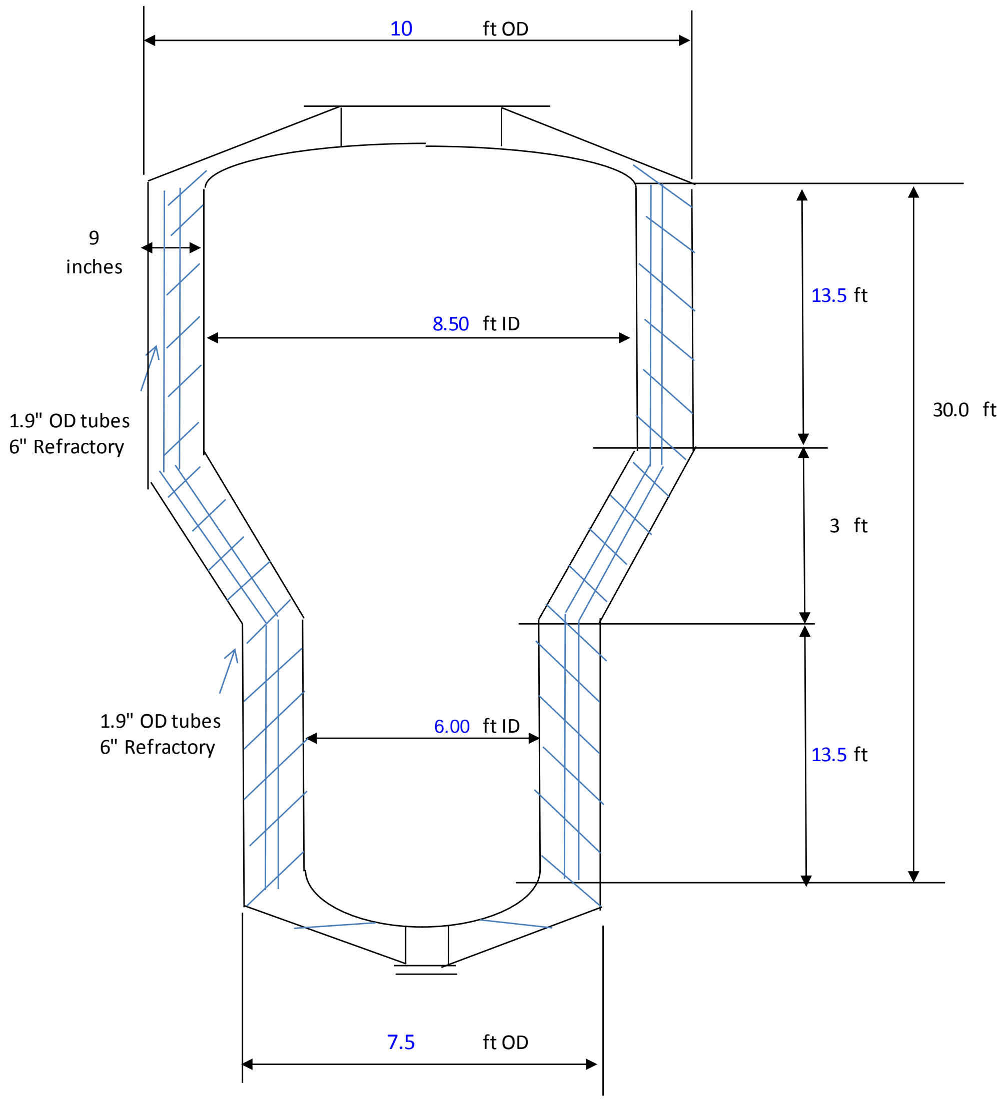
Two trains of ten GTI HMB gasifiers each are used to process a total of 15,527 tons/day of as-received Illinois No. 6 coal and 31,679 lbmol/h (289 MMSCFD) of natural gas (55% coal / 45% NG on HHV basis). These gasifiers operate at 515 psia. Coal is gasified with 95% oxygen from the ASU to produce syngas containing H_2, CO, CO_2, H_2O, NO_x, SO_x and other products of coal gasification. The gasifier wall is cooled by steam generation to create a protective ash layer over the refractory wall to maintain the gasification temperature at ~2,600°F. High carbon conversion (above 99%) is obtained in the gasifier, and the high temperature ensures that essentially no organic components heavier than methane are in the raw syngas. The insulation provided by the slag layer in the gasifier minimizes heat losses,

6.2.3. Gasifier Layout and Dimensions

Figure 6-1 is a conceptual layout of the GTI Hybrid Molten Bed Gasifier based on estimated dimensions provided by GTI. The layout and estimated dimensions formed the basis for estimating the cost of the HMB gasifier. Twenty of the HMB gasifiers are required.

HMB gasifier tests are currently performed by GTI and the test data will provide refinements to the gasifier dimensions and layout. The final technical details will be provided by GTI in a separate report.

Figure 6-1
GTI HMB Gasifier Conceptual Layout



This conceptual layout depicts the gasifier wall of built-in tube banks with a thin layer of castable refractory on the inside. A thin layer of frozen slag forms on the walls, protecting them from abrasion, a process demonstrated with many mineral melts in GTI submerged combustion melters. The dimensions of the gasifier are summarized in Table 6-1.

The cost of the GTI HMB gasifier for the GTI Case 1FT is based on the gasifier dimensions as shown. Allowance for burners, steam drums and circulating pumps, and slag removal are estimated and included in the overall gasifier cost.

Table 6-1
Case 1FT GTI HMB Gasifier Overall Dimensions

No. of Trains	2
No. of Gasifiers per Train	10
Gasifier Diameter (ID), ft (top)	8.5
Gasifier Diameter (ID),ft (bottom)	6
GTI Gasifier Height, ft (top)	13.5
GTI Gasifier Height, ft (bottom)	13.5
Refractory Thickness, inches, (top)	6
Refractory Thickness, inches, (bottom)	6
Steam Tube OD, inches	1.9
Gasifier Overall Dimensions:	
Shell Diameter (OD), ft (top)	10
Shell Diameter (OD), ft (bottom)	7.5
Swage Height, ft	3
Total Height, ft	30

6.2.4. Rectisol Acid Gas Removal (AGR)

The Rectisol AGR is designed for 90% CO_2 recovery and a product CO_2 purity of 95%. An acid gas stream is also produced with a minimum H_2S content of 25%. Additional CO_2 removal is required to achieve the < 10% CO_2 emissions target. A PC amine CO_2 removal unit is provided to remove > 85% of the CO_2 from the HRSG effluent gas.

Refer to section 5.2.9 for more detailed discussions of the Rectisol process.

6.2.5. Claus/TGTU Plant

Three Claus catalytic stages and a tail gas treating unit (TGTU) are required to achieve the sulfur recovery efficiency of 99.8%. Refer to section 5.2.10 for more detailed discussions of the Claus/TGTU process.

6.2.6. PSA Hydrogen Purification

A slipstream of treated syngas from the Rectisol Acid Gas Removal Unit is purified in the PSA unit to recover 99.99% hydrogen. The PSA unit is designed for 37 MMSCFD of purified hydrogen. The product hydrogen is mainly used in the naphtha hydrotreater and the wax hydrocracker units for the hydro-processing of the Fischer-Tropsch reactor liquid. A portion of the hydrogen is also used periodically for catalyst reduction in the FT unit. The residual gas from the PSA unit is sent to the LP fuel system

6.2.7. Fischer-Tropsch Synthesis

The FT synthesis unit is designed to operate at 413 psia and converts 212,571 lbmol/h of treated syngas into 36,611 BPD of FT diesel and 12,562 BPD of FT naphtha for the Case 1FT. Part of the tail gas generated by the FT synthesis and upgrading units are recycled to improve conversion and the remaining fuel gas is sent to the gas turbine for power generation. Only sufficient power is generated for the FT CNTL plant with net power export minimized. Refer to section 5.2.12 for detailed FT synthesis process descriptions.

6.2.8. CO_2 Compression and Dehydration

MP and LP flashed CO_2 from the Rectisol AGR plant are at the battery limit conditions of 42 and 19 psia respectively and at 80°F. Additional CO_2 from the PC amine CO_2 removal unit is also sent to CO_2 compression.

Refer to section 5.2.14 for process description.

6.2.9. Gas Turbine

The gas turbine generator selected is a GE SG6FA class turbine with a nominal ISO gross GT output of 95 MW. Nitrogen from the ASU is used for dilution to limit NO_x formation and to adjust the syngas LHV to 115-132 Btu/Scf. Inlet air is compressed to a pressure ratio of 15:1 for the GT combustion process. Hot combustion products are expanded in the gas turbine expander with an exhaust temperature of around 1,065°F. Two GTs are used for Case 1FT for a total of 190 MW GT output.

6.2.10. Steam Turbine and HRSG

The 1,065°F GT exhaust is cooled in the HRSG by generating HP, IP and LP steams for the steam turbines (ST) and process users. The cooled GT flue gas exits the HRSG at 200°F and is sent to a PC amine CO_2 removal unit before it is vented to the atmosphere through a stack. HP steam is used in the ST for power generation. LP exhaust steam from the last ST stage is condensed. The condensers operate at 0.698 psia with a corresponding condensing temperature of 90°F.

The condensates are collected and send 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 steam generators.

6.2.11. BOP

The BOP facilities are similar to the reference case. Refer to section 5.2.17 for detailed descriptions.

6.3. Case 1FT Performance

Table 6-2 shows the power production and auxiliary load breakdown of the Case 1FTGTI HMB gasification-based FT CNTL running on 55% coal and 45% NG co-feed. The FT CNTL design is based on minimizing the net power export. Sufficient FT tail gas is used in the gas turbines to generate power to satisfy the auxiliary loads. The balance of the FT tail gas is recycled to the gasifier to generate additional H_2/CO for the FT feed.

Table 6-2
Case 1FT Power Generation and Auxiliary Load Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case 1FT 55% Coal / 45% NG Direct Reforming
Gas Turbine Power	189,802
Steam Turbine Power	579,472
TOTAL POWER, kWe	**769,274**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling & Milling	25,050
Slag Handling	1,542
Natural Gas Compressors	25,299
Gasifier System	12,830
Air Separation Unit Main Air Compressor & Auxiliaries	243,756
Oxygen Compressor	139,076
CO2 Compressor	46,046
Boiler Feedwater Pumps	34,694
Condensate Pump	247
Circulating Water Pump	40,804
Ground Water Pumps	319
Cooling Tower Fans	13,718
Acid Gas Removal	47,250
Claus Plant/TGTU Auxiliaries	1,448
Sour Water Stripper	1,325
FT Power Requirement	50,229
PSA & TG Recycle	73
ATR w/Feed Pump	14,537
Miscellaneous Balance of Plant	64,774
TOTAL AUXILIARIES, kWe	**763,016**
NET POWER, kWe	**6,257**
CONSUMABLES	
As-Received Coal Feed, lb/hr	1,293,879
Natural Gas Feed, lbs/hr	501,204
Thermal Input, kWt	7,743,409
Condenser Duty, MMBtu/hr	1,616
Raw Water Withdrawal, gpm	32,072
Raw Water Consumption, gpm	29,617

6.4. Carbon Balance

Table 6-3 show the carbon balances for the Case 1FT GTI HMB gasifier-based FT CNTL. Carbon emission is based on the total stack emissions from the process heaters, cogen and SRU/TGTU. CO_2 is recovered from the AGR and the PC amine CO2 recovery units with the balance of carbons going to the FT products. The waste liquid effluent and ash byproducts contain a small amount of carbon.

Table 6-3
Case 1FT Carbon Balance

OVERALL CARBON BALANCE		
Case 1FT		
Case Description	55% Coal / 45% Coal Direct Reforming	
	lbs/hr	**%**
Carbon IN:		
Coal NG & Flux	1,221,239	99.97
Cogen Combustion Air	400	0.03
Total Carbon In	**1,221,639**	**100.00**
Carbon OUT:		
Process Htr Stack	40,056	3.28
Cogen Stack	18,109	1.48
SRU/TGTU Stack	35,523	2.91
Recovered C from AGR	532,908	43.62
Export FG	0	0.00
AGR Purge H2O	174	0.01
FT Waste H2O	7,570	0.62
Slag & Ash	4,490	0.37
Naphtha + Diesel Products	480,310	39.32
PC Amine CO2 Recovery	102,618	8.40
Convergence Error	-119	-0.01
Total Carbon Out	**1,221,639**	**100.00**
Carbon Emission:		
Total Carbon Emissions	101,432	7.67

6.5. Water balance

Water makeup and consumptions are shown in Table 6-4. The scrubber water demand is based on a maximum chlorides concentration in the scrubber purge water of 1,000 ppmw.

Table 6-4
Case 1FT Water Balance

Water Balance (GPM)	**Case 1FT** **55% Coal / 45% NG Co-Feed Direct Reforming**				
Water Usage by Area	Water Demand	Internal Recycle	Net Water Demand	Process Water Discharge	Raw Water Withdrawal
(GPM)					
Slag Handling	524		524	-	
SG Scrubber Makeup	6,486		6,486	(7,471)	
SWS	4,961	(4,961)	-		
SG Cooling Cond		(4,958)	(4,958)		
AGR/SRU/TGU Steam Cond		(4,420)	(4,420)	(232)	
FT Block			-	(1,311)	
BFW	21,858	(8,857)	13,000		
Blowdowns			-	(861)	
Fuel Gas Saturator	-		-		
Cooling Tower	18,937		18,937	(2,646)	
Potable Water + Contingency	47		47	7	
Total Water Usage	52,813	(23,196)	29,617	(12,514)	
Raw Water Treating					
RO/Demin System			(9,154)	(1,426)	10,533
Makeup Cooling Water Treating			(20,462)	(1,077)	21,539
Total Treated Water			(29,617)	(2,503)	32,072
Total	52,813	(23,196)	-	(15,017)	32,072

6.6. Capital Cost

6.6.1. Total Plant Cost

Table 6-5 shows the total plant cost (TPC) summary of Case 1FT. Account 4 includes the costs of the gasifier, natural gas compressors, syngas cooling and ASU.

Table 6-5
Case 1FT Total Plant Cost Summary

Total Plant Cost (June 2011)		GTI HMB Case 1FT
Acct. No.	**Item/Description**	**$MM**
1	COAL & SORBENT HANDLING	$ 80.53
2	COAL & SORBENT PREP & FEED	$ 423.37
3	FEEDWATER & MISC BOP SYSTEMS	$ 81.31
4	GASIFIER & ACCESSORIES	$ 1,960.94
5	GAS CLEANUP & PIPING	$ 1,122.87
5AA	FT SYNTHESIS AND PRODUCT UPGRADE	$ 1,111.24
5B.2	CO2 Compression & Drying	$ 94.60
6	COMBUSTION TURBINE/ACCESSORIES	$ 122.50
7	HRSG, PC AMINE UNIT, DUCTING & STACK	$ 119.31
8	STEAM TURBINE GENERATOR	$ 124.96
9	COOLING WATER SYSTEM	$ 88.77
10	ASH/SPENT SORBENT HANDLING SYS	$ 81.49
11	ACCESSORY ELECTRIC PLANT	$ 165.41
12	INSTRUMENTATION & CONTROL	$ 38.44
13	IMPROVEMENTS TO SITE	$ 50.60
14	BUILDINGS & STRUCTURES	$ 35.96
		$ 5,702.30

6.6.2. Operating Costs

Table 6-6 summarizes the operating costs for Case 1FT.

Table 6-6
Case 1FT Operating Cost Breakdown

OPERATING COSTS, 2011 $MM/yr	GTI HMB Case 1FT
FIXED OPERATING COSTS	
Annual Operting Labor Cost	$28.9
Maintainence Labor Cost	$51.2
Administration & Support Labor	$20.0
Property Taxes and Insurance	$114.0
TOTAL FIXED OPERATING COSTS	$214.2
VRIABLE OPERATING COSTS (@90% CF)	
NON-FUEL VARIABLE OPERATING COSTS	
Maintenance Material Cost	$96.7
Water	$12.7
Chemicals	
MU & WT Chemicals	$12.1
Other Chemicals & Catalysts	$15.8
Waste Disposal	$12.9
Power Credits	($2.9)
TOTAL NON_FUEL VARIABLE OPERATING COSTS	$147.4
FUEL (@90% CF)	
Coal	$349.9
Natural Gas	$506.1
TOTAL VARIABLE OPERATING COSTS	$1,003.4

6.7.Cost of Electricity

Table 6-7 shows a summary of the power output, CAPEX, OPEX, COP and cost of CO_2 capture for Case1 compared to the Reference Benchmark case.

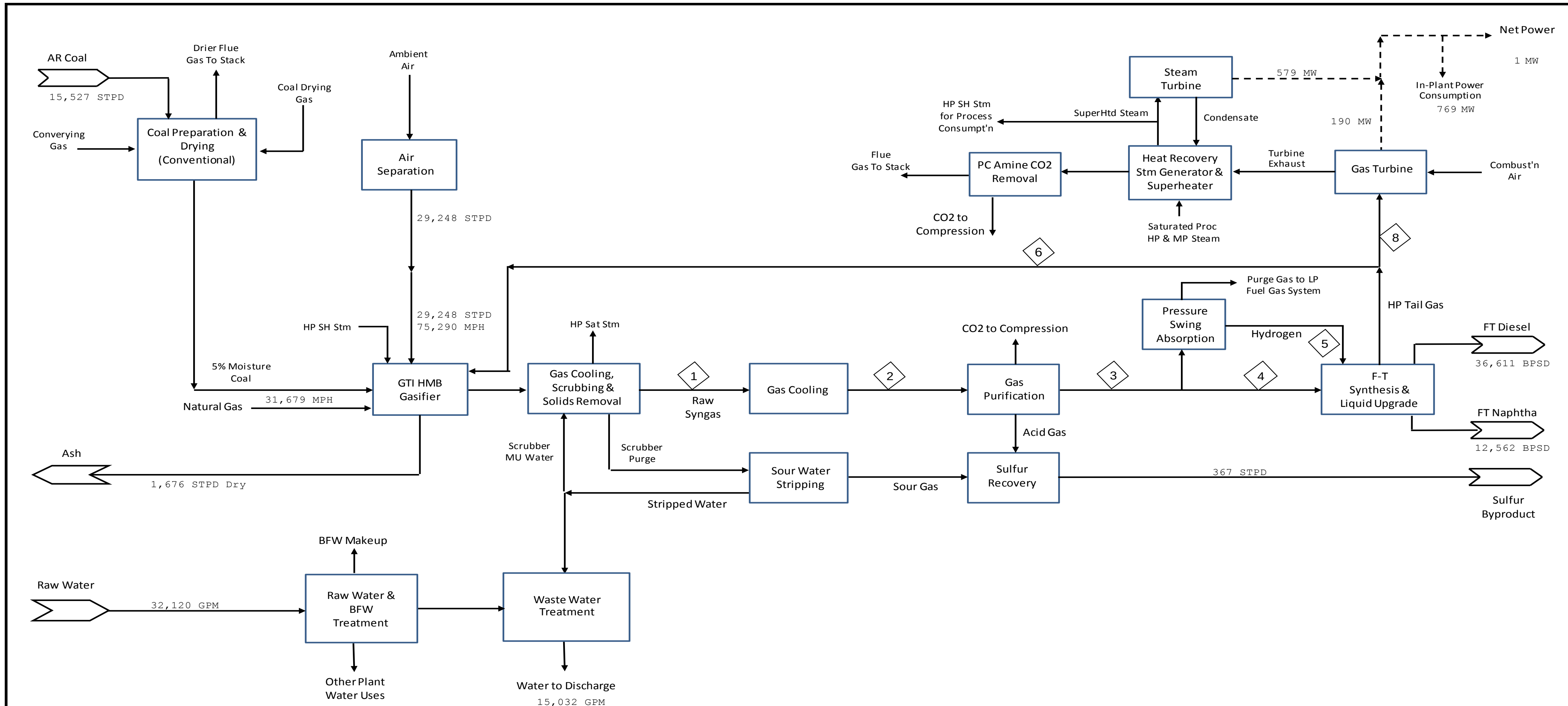
Table 6-7
Case 1FT Plant Performance and Economic Summary

	GTI HMB Case 1FT
CAPEX, $MM	
Total Installed Cost (TIC)	$4,152
Total Plant Cost (TPC)	$5,702
Total Overnight Cost (TOC)	$7,116
OPEX, $MM/yr (90% Capacity Factor Basis)	
Fixed Operating Cost (OC_{Fix})	$214
Variable Operating Cost Less Fuel (OC_{VAR})	$147
Fuel Cost (OC_{Fuel})	$856
Power Production, Mwe	
Gas Turbine	190
Steam Turbine	579
Auxiliary Power Consumption	763
Net Power Output	6
Power Generated, MWh/yr (MWH)	54,814
COP FT Diesel, excl CO2 TS&M, $/bbl FT diesel	186.7
COP FT Diesel, incl CO2 TS&M, $/bbl FT diesel	199.4
COP FT EPD, excl CO2 TS&M, $/bbl EPD	181.0
COP FT EPD, incl CO2 TS&M, $/bbl EPD	193.8
COP FT ECO, excl CO2 TS&M, $/bbl ECO	144.8
COP FT ECO, incl CO2 TS&M, $/bbl ECO	157.6
COP FT Naphtha, excl CO2 TS&M, $/bbl FT Naphtha	129.9
COP FT Naphtha, incl CO2 TS&M, $/bbl FT Naphtha	142.7

6.8.BFD & stream data

A block flow diagram with stream data for the GTI HMB Gasifier Case 1FT is shown below.

Figure 6-2
BFD for GTI HMB Gasifier Case 1FT



Steam No.	1	2	3	4	5	6	8		
Description	Raw Syngas	Cooled Syngas	Clean Syngas	Syngas to FT	H2 to Upgrade	Tail Gas to Upgrade	HP Tail Gas to GT		
Temperature,F	361	100	95	95	95	95	95		
Pressure, psia	465	450	430	413	393	340	340		
Flow, lb-mole/hr									
CO	79,703	79,703	79,703	75,240	0	15,457	5,627		
H2	118,163	118,163	118,163	112,310	3,971	22,226	6,602		
CO2	57,633	57,633	6,702	6,327	0	12,606	3,300		
CH4	2	2	2	2	0	1,460	365		
C2H6	0	0	0	0	0	162	40		
C3H8	0	0	0	0	0	318	80		
C4 +	0	0	0	0	0	360	90		
Ar + N2	19,794	19,794	19,794	18,693	40	14,942	4,157		
H2S	877	877	0	0	0	0	0		
H2O	138,178	583	0	0	0	24	6		
Total, lb-mole/hr	414,431	276,836	224,365	212,571	4,011	67,554	20,266		
Total, lb/hr	8,223,259	5,744,408	3,457,748	3,265,851	9,128	1,621,285	478,170		

Rev No.	Date	Revision	Engineer

Nexant, Inc.
San Francisco, California
DOE/NETL Advanced Gasification Technologies Program
GTI Hybrid Molten Bed Gasifier for Production of High Hydrogen Syngas Project
Overall Block Flow Diagram
Case 1FT
HMB Gasifier F-T CNTL Plant
55% Coal / 45% Natural Gas Co-Feed to HMB Gasifier with CO2 Capture

Nexant	Job No.	Drawing No.	Revision No.
	A02220	BFD - 001	0

7. CASE 2FT: GTI HMB GASIFIER FT CNTL PLANT, 81% ILLINOIS NO. 6 COAL / 19% NG FEED AND CO_2 CAPTURE (PARALLEL INDIRECT REFORMING)

7.1. Process Overview

Case 2FT is designed to generate the required H_2/CO ratio of 1.5 for the iron based FT synthesis using the parallel indirect reforming configuration. A simplified block flow diagram is shown in Figure 4-3.

This configuration utilizes an external steam methane catalytic reformer where natural gas and/or FT tail gas is reformed with steam. However, instead of returning the reformer syngas to the HMB gasifier, the relatively clean reformer syngas is cooled and processed for contaminant removal separately from the gasifier syngas. The reformer syngas contains very low concentration of particulates, mercury and sulfur species and is sent to the CO_2 removal section only in the Rectisol AGR. The gasifier syngas is cooled and cleaned to remove contaminants such as particulates, mercury, chlorides, ammonia and various sulfur species to minimize contaminants in the FT feed. The external steam methane reformer duty is provided by the 2,600°F syngas exiting the GTI HMB gasifier. The syngas exiting the reformer is at 1,500°F has a H_2/CO ratio of ~3. The HMB gasifier using an 81% coal / 19% NG co-feed and a reformer feed steam/C ratio of 1.3 mole/mole is capable of producing the required FT feed H_2/CO ratio of 1.5. The FT feed is a mix of the gasifier and reformer syngas.

As in Case 1FT, by generating the required FT feed syngas H_2/CO ratio of 1.5 in the HMB gasifier, Case 2FT eliminates the need for water gas shift reactors and reduces the cost of the FT CNTL plant. The FT CNTL plant is designed to produce a nominal 50,000 BPD of FT diesel and naphtha.

The iron based FT synthesis process configuration is similar to the reference Shell case except the steam methane reformer is used instead of the autothermal reformer (ATR) to convert the hydrogen, CO, and hydrocarbons in the FT tail gas to syngas. The FT tail gas is recycled to the steam methane reformer where it replaces and reduces the natural gas consumption in the steam methane reformer.

The steam, power generation section and the balance of plant to support the gasification and FT synthesis / products upgrading systems consists of the same units as the Reference case.

7.2. Process Description

The process descriptions for the various Case 2FT subsystems are identical to those described for the Shell Reference FT CNTL case in Section 5.2, except for the dual-fueled gasification section, which runs on a combination of natural gas and coal, as described in Section 7.2.1 below and the addition of the steam methane reformer, as described in Section 7.2.4, which converts the natural gas feed into reformed syngas to be fired into the HMB gasifier.

7.2.1. ASU

ASU for Case 2FT is similar to the reference case ASU. Case 2FT requires 18,124 tons/day of 95mol% oxygen for the GTI HMB gasifier.

7.2.2. GTI HMB Gasification

Two trains of five GTI HMB gasifiers each are used to process a total of 20,702 tons/day of as-received Illinois No. 6 coal and 12,000 lbmol/h (109 MMSCFD) of natural gas (81% coal / 19% NG on HHV basis). These gasifiers operate at 515 psia. Coal is gasified with 95% oxygen from the ASU to produce syngas containing H_2, CO, CO_2, H_2O, NO_x, SO_x and other products of coal gasification. The gasifier wall is cooled by steam generation to create a protective ash layer over the refractory wall to maintain the gasification temperature at ~2,600°F. High carbon conversion (above 99%) is obtained in the gasifier, and the high temperature ensures that essentially no organic components heavier than methane are in the raw syngas. The insulation provided by the slag layer in the gasifier minimizes heat losses.

7.2.3. Gasifier Layout and Dimensions

The layout and estimated dimensions for each of the eight HMB gasifiers for Case 2FT are the same as Case 1FT and are shown in Figure 6-1.

HMB gasifier tests are currently performed by GTI and the test data will provide refinements to the gasifier dimensions and layout. The final technical details will be provided by GTI in a separate report.

7.2.4. Steam Methane Reformer

A conceptual design of the syngas heated steam methane reformer is shown in Figure 7-2. GTI will provide additional details into the technical development and operational aspects of the HMB steam methane reformer in another report.

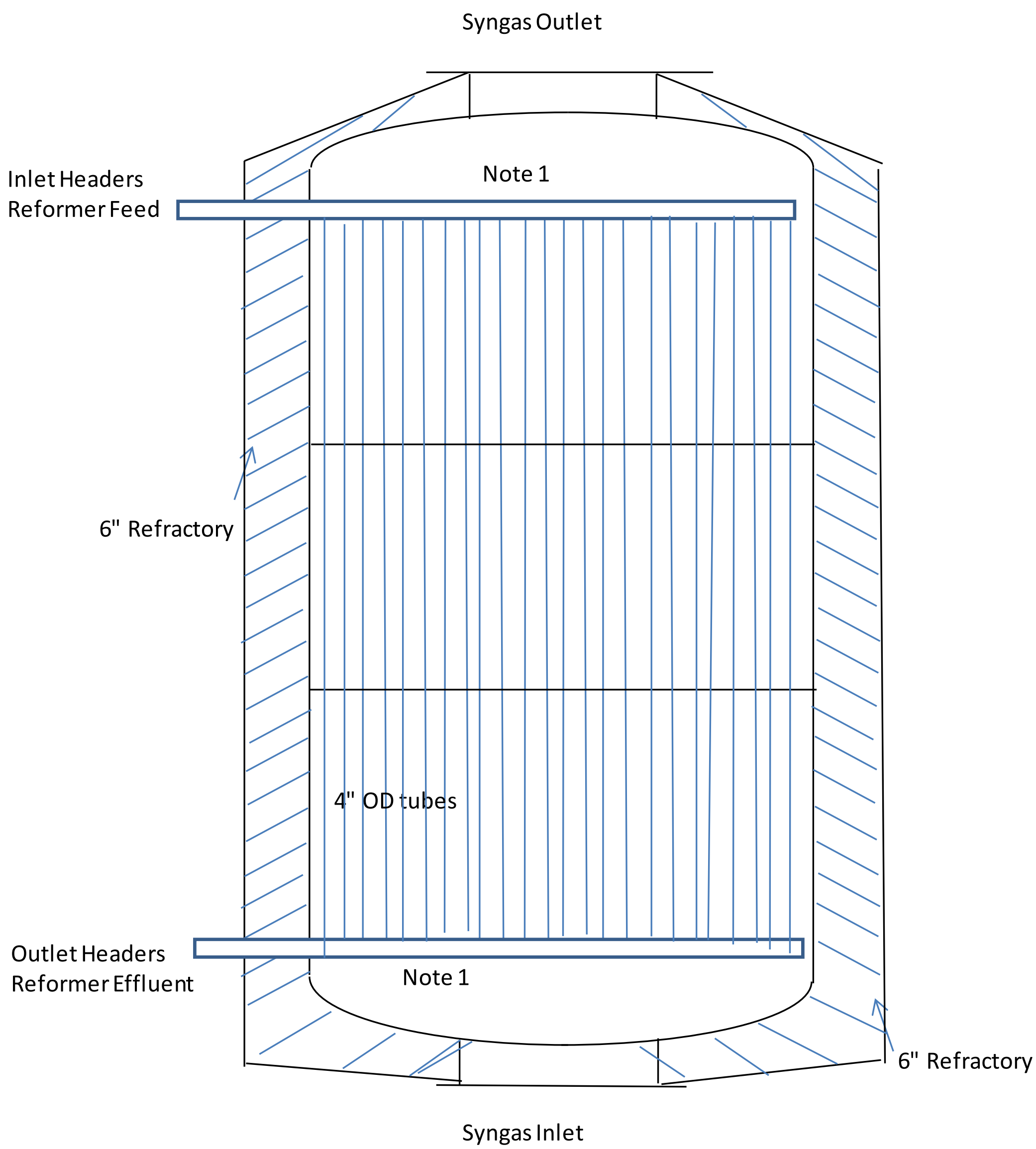
The key features of the steam methane reformer are:

- Shell Side – refractory lined
  - Syngas on shell side; Inlet temperature at ~2,600°F, Outlet temperature at ~ 1,800°F
- Tube Side – catalyst filled
  - Steam and natural gas feed; Inlet temperature at ~1,085°F, Outlet temperature at ~ 1,500°F
- Reformer duty of approximately 178 MMBtu/h per reformer, with a total reformer duty of 2,124 MMBtu/h for 12 reformers

The cost of the steam methane reformer for Case 2FT is based on the reformer concept as shown in the conceptual layout in Figure 7-1. The reformer dimensions are estimated from the tube heat exchange surface and catalyst volume requirements based on reforming duty.

HK40 material is assumed for the high temperature reforming tube material of construction to provide additional contingency for the reformer cost. Traditional steam methane reformer tube materials of construction are HK40 or equivalent, which are high Cr and Ni alloys for high temperature service. The reformer vessel wall is assumed to be stainless steel construction with 6” of refractory.

Figure 7-1
Syngas Heated Steam Methane Reformer Conceptual Layout



Note 1) Tubes are connected to the headers through pigtails to allow for tube thermal exxpansion

7.2.5. Syngas Cooling & Particulate Filters

The Case 2FT syngas cooling /heat integration is optimized to provide reforming duty for the steam methane reformer, preheating duties for natural gas and reformer steam feeds and high pressure steam generation/superheat. The primary goal is to provide sufficient duty for the steam methane reformer and feed preheat requirements. The remaining cooling duty is used for high pressure steam generation and superheating. Hot 2,600°F syngas generated by the HMB gasifier is heat exchanged with the steam methane reformer. The cooled syngas exits the reformer and is cooled further by providing preheating duties for the natural gas feed (900°F) and reformer steam (1,200°F). After feed preheating, the raw syngas is cooled to 685°F by high pressure steam generation before entering the ceramic particulate filters and cyclones. Any remaining particulate matters in the syngas will be removed by these particulate filters and cyclones.

7.2.6. Rectisol Acid Gas Removal (AGR)

The Rectisol AGR is designed for 90% CO_2 recovery and a product CO_2 purity of 95%. An acid gas stream is also produced with a minimum H_2S content of 25%. Additional CO_2 removal is required to achieve the < 10% CO_2 emissions target. A PC amine CO_2 removal unit is provided to remove > 85% of the CO_2 from the HRSG effluent gas.

Refer to section 5.2.9 for more detailed discussions of the Rectisol process.

7.2.7. Claus Plant

Three Claus catalytic stages and a tail gas treating unit (TGTU) are required to achieve the sulfur recovery efficiency of 99.8%. Refer to section 5.2.10 for more detailed discussions of the Claus/TGTU process.

7.2.8. PSA Hydrogen Purification

A slipstream of treated syngas from the Rectisol Acid Gas Removal Unit is purified in the PSA unit to recover 99.99% hydrogen. The PSA unit is designed for 36 MMSCFD of purified hydrogen. The product hydrogen is mainly used in the naphtha hydrotreater and the wax hydrocracker units for the hydro-processing of the Fischer-Tropsch reactor liquid. A portion of the hydrogen is also used periodically for catalyst reduction in the FT unit. The residual gas from the PSA unit is sent to the LP fuel system

7.2.9. Fischer-Tropsch Synthesis

The FT synthesis unit is designed to operate at 413 psia and convert 223,101 lbmol/h of treated syngas into 37,193 BPD of FT diesel and 12,762 BPD of FT naphtha for the Case 1FT. Part of the tail gas generated by the FT synthesis and upgrading units are recycled to improve conversion and the remaining fuel gas is sent to the gas turbine for power generation. Only sufficient power is generated for the FT CNTL plant with net power export minimized.
Refer to section 5.2.12 for detailed FT synthesis process descriptions.

7.2.10. CO2 Compression and Dehydration

MP and LP flashed CO_2 from the Rectisol AGR plant are at the battery limit conditions of 42 and 19 psia respectively and at 80°F. Additional CO_2 from the PC amine CO_2 removal unit is also sent to CO_2 compression.

Refer to section 5.2.14 for process description.

7.2.11. Gas Turbine

The gas turbine generator selected is a GE SG6FA class turbine with a nominal ISO gross GT output of 95 MW. Nitrogen from the ASU is used for dilution to limit NO_x formation and to adjust the syngas LHV to 115-132 Btu/Scf. Inlet air is compressed to a pressure ratio of 15:1 for the GT combustion process. Hot combustion products are expanded in the gas turbine expander with an exhaust temperature of around 1,014°F. Five GTs are used for Case 2FT for a total of 414 MW GT output.

7.2.12. Steam Turbine and HRSG

The 1,014°F GT exhaust is cooled in the HRSG by generating HP, IP and LP steams for the steam turbines (ST) and process users. The cooled GT flue gas exits the HRSG at 208°F and is sent to a PC amine CO_2 removal unit before it is vented to the atmosphere through a stack. HP steam is used in the ST for power generation. LP exhaust steam from the last ST stage is condensed. The condensers operate at 0.698 psia with a corresponding condensing temperature of 90°F.

The condensates are collected and send 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 steam generators.

7.2.13. BOP

The BOP facilities are similar to the reference case. Refer to section 5.2.17 for detailed descriptions.

7.3. Case 2FT Performance

Table 7-1 shows the power production and auxiliary load breakdown of the Case 2FT co-fired GTI HMB gasification-based FT CNTL running on 55% coal feed/45% natural gas feed. Sufficient FT tail gas is used in the gas turbines to generate power to satisfy the auxiliary loads. The balance of the FT tail gas is converted in the steam methane reformer to generate additional H_2/CO for the FT feed. The recycled FT tail gas replaces some of the natural gas feed to the reformer.

Table 7-1
Case 2FT Power Generation and Auxiliary Load Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	**Case 2FT 81% Coal / 19% NG Parallel Indirect Reforming**
Gas Turbine Power	414,195
Steam Turbine Power	160,567
TOTAL POWER, kWe	**574,762**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling & Milling	32,912
Slag Handling	2,056
Natural Gas Compressors	8,283
Gasifier System	11,437
Air Separation Unit Main Air Compressor & Auxiliaries	151,044
Oxygen Compressor	86,179
CO2 Compressor	45,435
Boiler Feedwater Pumps	32,316
Condensate Pump	69
Circulating Water Pump	26,512
Ground Water Pumps	560
Cooling Tower Fans	7,273
Acid Gas Removal	25,556
Claus Plant/TGTU Auxiliaries	1,931
Sour Water Stripper	611
FT Power Requirement	51,027
PSA & TG Recycle	73
ATR w/Feed Pump	19,912
Miscellaneous Balance of Plant	70,545
TOTAL AUXILIARIES, kWe	**573,730**
NET POWER, kWe	**1,031**
CONSUMABLES	
As-Received Coal Feed, lb/hr	1,725,173
Natural Gas Feed, lbs/hr	189,859
Thermal Input, kWt	7,155,818
Condenser Duty, MMBtu/hr	448
Raw Water Withdrawal, gpm	34,280
Raw Water Consumption, gpm	30,920

7.4. Carbon Balance

Table 7-2 show the carbon balances for the Case 2FT co-fired GTI HMB gasifier-based FT CNTL. Carbon emission is based on the total stack emissions from the process heaters, cogen and SRU/TGTU. CO_2 is recovered from the AGR and the PC amine CO2 recovery units with the balance of carbons going to the FT products. The waste liquid effluent and ash byproducts contain a small amount of carbon.

Table 7-2
Case 2FT Carbon Balance

OVERALL CARBON BALANCE		
Case 2FT		
Case Description	81% Coal / 19% NG Indirect External Parallel Reforming	
	lbs/hr	**%**
Carbon IN:		
Coal, NG & Flux	1,249,879	99.92
Cogen Combustion Air	1,051	0.08
Total Carbon In	**1,250,930**	**100.00**
Carbon OUT:		
Process Htr Stack	39,621	3.17
Cogen Stack	33,135	2.65
SRU/TGTU Stack	47,364	3.79
Recovered C from AGR	439,326	35.12
Export FG	0	0.00
AGR Purge H2O	91	0.01
FT Waste H2O	7,690	0.61
Slag & Ash	5,986	0.48
Naphtha + Diesel Products	489,979	39.17
PC Amine CO2 (carbon) Recovery	187,766	15.01
Convergence Error	-29	0.00
Total Carbon Out	**1,250,930**	**100.00**
Carbon Emission:		
Total Carbon Emissions	127,902	9.60

7.5. Water balance

Water makeup and consumptions are shown in Table 6-4. The scrubber water demand is based on a maximum chlorides concentration in the scrubber purge water of 1,000 ppmw.

Table 7-3
Case 2FT Water Balance

Water Balance (GPM)	**Case 2FT** **81% Coal / 19% NG Co-Feed Indirect Parallel Reforming**				
Water Usage by Area	Water Demand	Internal Recycle	Net Water Demand	Process Water Discharge	Raw Water Withdrawal
(GPM)					
Slag Handling	699		699	-	
SG Scrubber Makeup	9,969		9,969	(9,962)	
SWS	2,311	(2,311)	-		
SG Cooling Cond		(2,309)	(2,309)		
AGR/SRU/TGU Steam Cond		(5,678)	(5,678)	(196)	
FT Block			-	(1,332)	
BFW	20,337	(2,473)	17,865		
Blowdowns			-	(476)	
Fuel Gas Saturator	-		-		
Cooling Tower	10,326		10,326	(1,141)	
Potable Water + Contingency	48		48	8	
Total Water Usage	43,691	(12,771)	30,920	(13,099)	
Raw Water Treating					
RO/Demin System			(12,935)	(2,461)	15,349
Makeup Cooling Water Treating			(17,984)	(947)	18,931
Total Treated Water			(30,920)	(3,408)	34,280
Total	43,691	(12,771)	-	(16,507)	34,280

7.6. Capital Cost

7.6.1. Total Plant Cost

Table 7-4 shows the total plant cost (TPC) summary of Case 2FT. Account 4 includes the cost of the gasifier, steam methane reformer, natural gas compressors, syngas cooling and ASU.

Table 7-4
Case 2FT Total Plant Cost Summary

Total Plant Cost (June 2011)		GTI HMB Case 2FT	
Acct. No.	**Item/Description**	**$MM**	
1	COAL & SORBENT HANDLING	$	96.25
2	COAL & SORBENT PREP & FEED	$	511.90
3	FEEDWATER & MISC BOP SYSTEMS	$	67.55
4	GASIFIER & ACCESSORIES	$	1,758.22
5	GAS CLEANUP & PIPING	$	1,187.55
5AA	FT SYNTHESIS AND PRODUCT UPGRADE	$	1,041.19
5B.2	CO2 Compression & Drying	$	77.69
6	COMBUSTION TURBINE/ACCESSORIES	$	194.65
7	HRSG, PC AMINE UNIT, DUCTING & STACK	$	163.49
8	STEAM TURBINE GENERATOR	$	50.57
9	COOLING WATER SYSTEM	$	60.40
10	ASH/SPENT SORBENT HANDLING SYS	$	97.54
11	ACCESSORY ELECTRIC PLANT	$	140.78
12	INSTRUMENTATION & CONTROL	$	36.72
13	IMPROVEMENTS TO SITE	$	50.38
14	BUILDINGS & STRUCTURES	$	35.85
		$	5,570.70

7.7. Operating Costs

Table 7-5 summarizes the operating costs for GTI HMB Case 2FT.

Table 7-5
Case 2FT Operating Cost Breakdown

OPERATING COSTS, 2011 $MM/yr	GTI HMB Case 2FT
FIXED OPERATING COSTS	
Annual Operting Labor Cost	$28.9
Maintainence Labor Cost	$50.0
Administration & Support Labor	$19.7
Property Taxes and Insurance	$111.4
TOTAL FIXED OPERATING COSTS	$210.1
VRIABLE OPERATING COSTS (@90% CF)	
NON-FUEL VARIABLE OPERATING COSTS	
Maintenance Material Cost	$94.5
Water	$13.6
Chemicals	
MU & WT Chemicals	$12.9
Other Chemicals & Catalysts	$14.0
Waste Disposal	$17.3
Power Credits	($0.5)
TOTAL NON_FUEL VARIABLE OPERATING COSTS	$151.8
FUEL (@90% CF)	
Coal	$466.5
Natural Gas	$191.7
TOTAL VARIABLE OPERATING COSTS	$810.1

7.8.Cost of Electricity

Table 7-6 shows a summary of the power output, CAPEX, OPEX, COP and cost of CO_2 capture for Case 2FT.

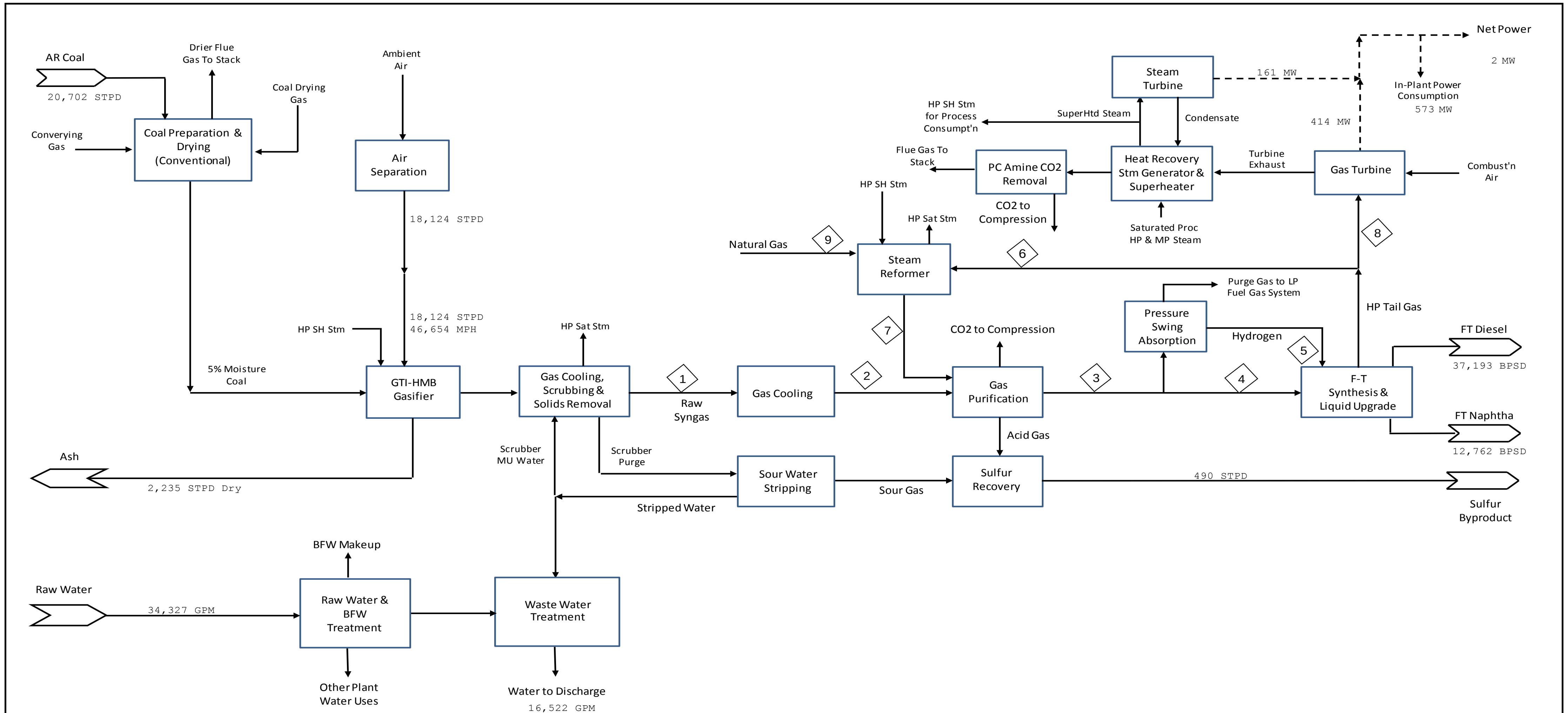
Table 7-6
Case 2FT Plant Performance and Economic Summary

	GTI HMB Case 2FT
CAPEX, $MM	
Total Installed Cost (TIC)	$3,926
Total Plant Cost (TPC)	$5,571
Total Overnight Cost (TOC)	$6,920
OPEX, $MM/yr (90% Capacity Factor Basis)	
Fixed Operating Cost (OC_{Fix})	$210
Variable Operating Cost Less Fuel (OC_{VAR})	$152
Fuel Cost (OC_{Fuel})	$658
Power Production, Mwe	
Gas Turbine	414
Steam Turbine	161
Auxiliary Power Consumption	574
Net Power Output	1
Power Generated, MWh/yr (MWH)	9,036
COP FT Diesel, excl CO2 TS&M, $/bbl FT diesel	167.1
COP FT Diesel, incl CO2 TS&M, $/bbl FT diesel	176.0
COP FT EPD, excl CO2 TS&M, $/bbl EPD	162.0
COP FT EPD, incl CO2 TS&M, $/bbl EPD	170.9
COP FT ECO, excl CO2 TS&M, $/bbl ECO	129.6
COP FT ECO, incl CO2 TS&M, $/bbl ECO	138.5
COP FT Naphtha, excl CO2 TS&M, $/bbl FT Naphtha	116.3
COP FT Naphtha, incl CO2 TS&M, $/bbl FT Naphtha	125.2

7.9 BFD & stream data

A block flow diagram with stream data for the GTI HMB Gasifier Case 2FT is shown below.

Figure 7-2
BFD for GTI HMB Gasifier Case 2FT



Steam No.	1	2	3	4	5	6	7	8	9
Description	Raw Syngas	Shifted Syngas	Clean Syngas	Syngas to FT	H2 to Upgrade	Tail Gas to Upgrade	Upgraded Gas to AGR	HP Tail Gas to GT	Natral Gas to Steam Reformer
Temperature,F	348	100	95	95	95	95	95	95	95
Pressure, psia	460	445	425	413	393	340	445	340	502
Flow, lb-mole/hr									
CO	61,813	61,813	81,769	77,283	0	14,906	19,956	7,672	0
H2	61,390	61,390	119,986	114,096	3,949	20,548	58,596	8,911	0
CO2	33,818	33,818	7,029	6,644	0	11,811	18,309	4,594	120
CH4	3	3	16,023	15,144	0	12,373	16,021	5,035	11,172
C2H6	0	0	0	0	0	150	0	56	384
C3H8	0	0	0	0	0	294	0	110	84
C4 +	0	0	0	0	0	332	0	124	48
Ar + N2	3,090	3,090	10,503	9,934	40	7,221	7,413	2,949	192
H2S	1,170	1,170	0	0	0	0	0	0	0
H2O	64,416	345	0	0	0	22	248	8	0
Total, lb-mole/hr	225,806	161,734	235,311	223,101	3,989	67,656	120,543	29,458	12,000
Total, lb/hr	4,657,704	3,503,414	3,459,582	3,271,394	9,079	1,465,054	1,997,836	632,075	207,933

Rev No.	Date	Revision	Engineer

Nexant, Inc.
San Francisco, California
DOE/NETL Advanced Gasification Technologies Program
GTI Hybrid Molten Bed Gasifier for Production of High Hydrogen Syngas Project
Overall Block Flow Diagram
Case 2FT
HMB Gasifier F-T CNTL Plant
81% Coal / 19% Natural Gas Co-Feed with CO2 Capture (Parallel Indirect Reforming)

Job No.	Drawing No.	Revision No.
A02220	BFD - 002	0

8. CASE 3: GTI HMB GASIFIER FT CNTL PLANT WITH 55% ILLINOIS NO. 6 COAL / 45% NG FEED, REFORMER AND CO2 CAPTURE

8.1. Process Overview

Case 3FT is designed to generate the required H_2/CO ratio of 1.5 for the iron based FT synthesis using the series indirect reforming configuration. A simplified block flow diagram is shown in Figure 4-4.

This configuration utilizes an external steam methane catalytic reformer where natural gas and/or FT tail gas is reformed with steam. The reformer duty is provided by the 2,600°F syngas exiting the GTI HMB gasifier. Case 3FT differs from Case 2FT in that the reformer syngas is returned to the HMB gasifier through the dual feed gasifier burners carrying with it the recuperated heat from the gasifier syngas. The heat recuperation improves plant efficiency and is possible due to the unique feature of the GTI dual feed gasifier for handling gaseous feed. The syngas exiting the reformer is at 1,500°F has a H_2/CO ratio of ~3. The HMB gasifier using a 55% coal / 45% NG co-feed and a reformer feed steam/C ratio of 1.4 mole/mole is capable of producing the required FT feed H_2/CO ratio of 1.5.

The gasifier syngas is cooled and cleaned to remove contaminants such as particulates, mercury, chlorides, ammonia and various sulfur species to minimize contaminants in the FT feed.

As in Cases 1FT and 2FT, by generating the required FT feed syngas H_2/CO ratio of 1.5 in the HMB gasifier, Case 3FT eliminates the need for water gas shift reactors and reduces the cost of the FT CNTL plant. The FT CNTL plant is designed to produce a nominal 50,000 BPD of FT diesel and naphtha.

The iron based FT synthesis process configuration is similar to the reference Shell case except the steam methane reformer is used instead of the autothermal reformer (ATR) to convert the hydrogen, CO, and hydrocarbons in the FT tail gas to syngas. The FT tail gas is recycled to the steam methane reformer where it replaces and reduces the natural gas consumption in the steam methane reformer.

The steam, power generation section and the balance of plant to support the gasification and FT synthesis / products upgrading systems consists of the same units as the Reference case.

8.2. Process Description

The process descriptions for the various Case 3FT subsystems are identical to those described for Case 1FT in Section 6.2, except for the addition of the steam methane reformer, as described in Section 8.2.3, which converts the natural gas feed into reformed syngas to be fired into the HMB gasifier.

8.2.1. ASU

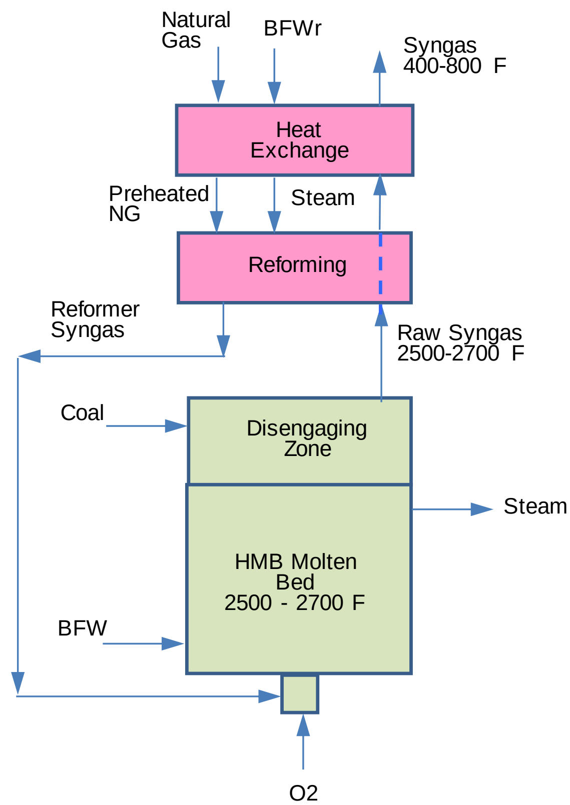
ASU for Case 2FT is similar to the reference case ASU. Case 3FT requires 19,599 tons/day of 95mol% oxygen for the GTI HMB gasifier.

8.2.2. GTI HMB Gasification

Two trains of three GTI HMB gasifiers each are used to process a total of 13,462 tons/day of as-received Illinois No. 6 coal and 26,826 lbmol/h (244 MMSCFD) of natural gas (55% coal / 45% NG on HHV basis). These gasifiers operate at 515 psia. Coal is gasified with 95% oxygen from the ASU to produce syngas containing H_2, CO, CO_2, H_2O, NO_x, SO_x and other products of coal gasification. The gasifier wall is cooled by steam generation to create a protective ash layer over the refractory wall to maintain the gasification temperature at ~2,600°F. High carbon conversion (above 99%) is obtained in the gasifier, and the high temperature ensures that essentially no organic components heavier than methane are in the raw syngas. The insulation provided by the slag layer in the gasifier minimizes heat losses.

In the HMB gasifier for Case 3FT, in place of natural gas, reformed syngas and oxygen are fired under partial oxidation conditions upward into a bed of molten coal slag. The heat and gases generated drive the gasification process. Evaporative cooling walls generate steam for the external steam methane reformer to increase process efficiency. Optimization of the coal, natural gas, oxygen, and steam and their ratios to the gasifier/steam methane reformer generates a syngas with H_2/CO ratio of 1.5. A simple schematic of the Case 3FT gasifier/steam methane reformer system is shown in Figure 8-1 below

Figure 8-1
Case 3FT Gasifier/Steam Methane Reformer System Schematic



8.2.3. Gasifier Layout and Dimensions

The Case 3FT Gasifier layout and dimensions are the same as in Case 1FT (see Figure 6-1). The layout and estimated dimensions formed the basis for the cost estimation for the HMB gasifier.

8.2.4. Steam Methane Reformer

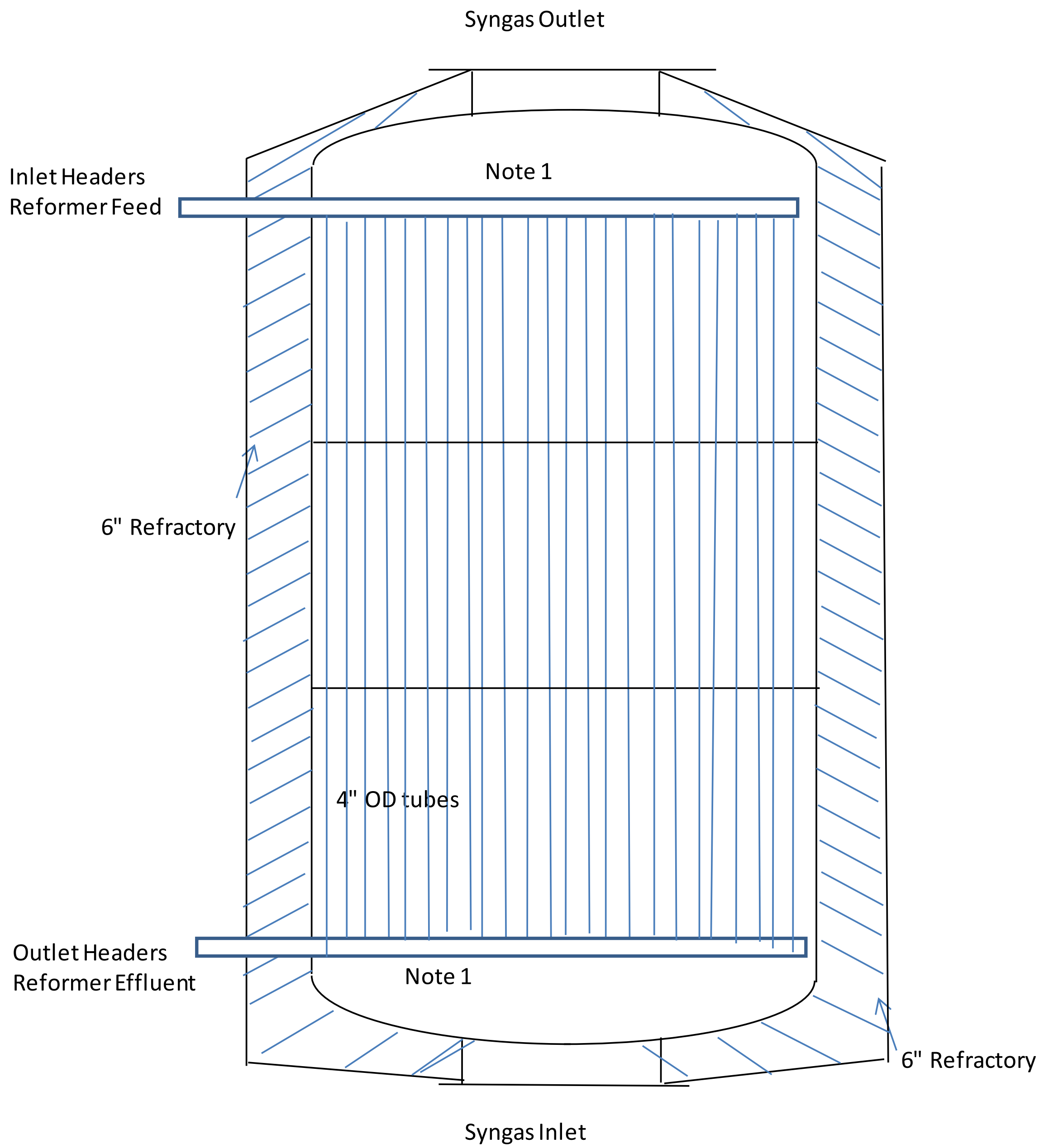
A conceptual design of the syngas heated steam methane reformer is shown in Figure 8-2. GTI will provide additional details into the technical development and operational aspects of the HMB steam methane reformer in another report.

The key features of the steam methane reformer are:

- Shell Side – refractory lined
  - Syngas on shell side; Inlet temperature at ~2,600°F, Outlet temperature at ~ 1,800°F
- Tube Side – catalyst filled
  - Steam and natural gas feed; Inlet temperature at ~1,085°F, Outlet temperature at ~ 1,500°F
- Reformer duty of approximately 176 MMBtu/h per reformer, with a total reformer duty of 2,461 MMBtu/h for 14 reformers

The cost of the steam methane reformer for Case 3FT is based on the reformer concept as shown in the conceptual layout in Figure 8-2. The reformer dimensions are estimated from the tube heat exchange surface and catalyst volume requirements based on reforming duty. HK40 material is assumed for the high temperature reforming tube material of construction to provide additional contingency for the reformer cost. Traditional steam methane reformer tube materials of construction are HK40 or equivalent, which are high Cr and Ni alloys for high temperature service. The reformer vessel wall is assumed to be stainless steel construction with 6" of refractory.

Figure 8-2
Syngas Heated Steam Methane Reformer Conceptual Layout



Note 1) Tubes are connected to the headers through pigtails to allow for tube thermal exxpansion

8.2.5. Syngas Cooling & Particulate Filters

The Case 3FT syngas cooling /heat integration is optimized to provide reforming duty for the steam methane reformer, preheating duties for natural gas and reformer steam feeds and high pressure steam generation/superheat. The primary goal is to provide sufficient duty for the steam methane reformer and feed preheat requirements. The remaining cooling duty is used for high pressure steam generation and superheating. Hot 2,600°F syngas generated by the HMB gasifier is heat exchanged with the steam methane reformer. The cooled syngas exits the reformer at ~1,800°F, hot enough to provide preheating duties for the natural gas feed (900°F) and reformer steam (1,200°F). After feed preheating, the raw syngas is cooled to 685°F by high pressure steam generation before entering the ceramic particulate filters and cyclones. Any remaining particulate matters in the syngas will be removed by these particulate filters and cyclones.

8.2.6. Rectisol Acid Gas Removal (AGR)

The Rectisol AGR is designed for 90% CO_2 recovery and a product CO_2 purity of 95%. An acid gas stream is also produced with a minimum H_2S content of 25%. %. Additional CO_2 removal is required to achieve the < 10% CO_2 emissions target. A PC amine CO_2 removal unit is provided to remove > 85% of the CO_2 from the HRSG effluent gas.

Refer to section 5.2.9 for more detailed discussions of the Rectisol process.

8.2.7. Claus/TGTU Plant

Three Claus catalytic stages and a tail gas treating unit (TGTU) are required to achieve the sulfur recovery efficiency of 99.8%. Refer to section 5.2.10 for more detailed discussions of the Claus/TGTU process.

8.2.8. PSA Hydrogen Purification

A slipstream of treated syngas from the Rectisol Acid Gas Removal Unit is purified in the PSA unit to recover 99.99% hydrogen. The PSA unit is designed for 36 MMSCFD of purified hydrogen. The product hydrogen is mainly used in the naphtha hydrotreater and the wax hydrocracker units for the hydro-processing of the Fischer-Tropsch reactor liquid. A portion of the hydrogen is also used periodically for catalyst reduction in the FT unit. The residual gas from the PSA unit is sent to the LP fuel system

8.2.9. Fischer-Tropsch Synthesis

The FT synthesis unit is designed to operate at 413 psia and convert 207,433 lbmol/h of treated syngas into 37,335 BPD of FT diesel and 12,811 BPD of FT naphtha for the Case 3FT. Part of the tail gas generated by the FT synthesis and upgrading units are recycled to improve conversion and the remaining fuel gas is sent to the gas turbine for power generation. Only sufficient power is generated for the FT CNTL plant with net power export minimized.
Refer to section 5.2.12 for detailed FT synthesis process descriptions.

8.2.10. CO_2 Compression and Dehydration

MP and LP flashed CO_2 from the Rectisol AGR plant are at the battery limit conditions of 42 and 19 psia respectively and at 80°F. Additional CO_2 from the PC amine CO_2 removal unit is also sent to CO_2 compression.

Refer to section 5.2.14 for process description.

8.2.11. Gas Turbine

The gas turbine generator selected is a GE SG6FA class turbine with a nominal ISO gross GT output of 95 MW. Nitrogen from the ASU is used for dilution to limit NO_x formation and to adjust the syngas LHV to 115-132 Btu/Scf. Inlet air is compressed to a pressure ratio of 15:1 for the GT combustion process. Hot combustion products are expanded in the gas turbine expander with an exhaust temperature of around 1,056°F. Three GTs are used for Case 3FT for a total of 256 MW GT output.

8.2.12. Steam Turbine and HRSG

The 1,056°F GT exhaust is cooled in the HRSG by generating HP, IP and LP steams for the steam turbines (ST) and process users. The cooled GT flue gas exits the HRSG at 208°F and is vented to the atmosphere through a stack. HP steam is used in the ST for power generation. LP exhaust steam from the last ST stage is condensed. The condensers operate at 0.698 psia with a corresponding condensing temperature of 90°F.

The condensates are collected and send 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 steam generators.

8.2.13. BOP

The BOP facilities are similar to the reference case. Refer to section 5.2.17 for detailed descriptions.

8.3.Performance

Table 8-1 shows the power production and auxiliary load breakdown for Case 3FT. Sufficient FT tail gas is used in the gas turbines to generate power to satisfy the auxiliary loads. The balance of the FT tail gas is converted in the steam methane reformer to generate additional H_2/CO for the FT feed. The recycled FT tail gas replaces some of the natural gas feed to the reformer.

Table 8-1
Case 3FT Power Generation and Auxiliary Load Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	**Case 3FT 55% Coal / 45% NG Series Indirect Reforming**
Gas Turbine Power	271,051
Steam Turbine Power	303,068
TOTAL POWER, kWe	**574,119**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling & Milling	21,718
Slag Handling	1,337
Natural Gas Compressors	10,634
Gasifier System	10,232
Air Separation Unit Main Air Compressor & Auxiliaries	163,342
Oxygen Compressor	93,195
CO2 Compressor	33,992
Boiler Feedwater Pumps	27,059
Condensate Pump	146
Circulating Water Pump	28,906
Ground Water Pumps	298
Cooling Tower Fans	9,174
Acid Gas Removal	30,319
Claus Plant/TGTU Auxiliaries	1,256
Sour Water Stripper	903
FT Power Requirement	51,223
PSA & TG Recycle	72
ATR w/Feed Pump	17,801
Miscellaneous Balance of Plant	64,662
TOTAL AUXILIARIES, kWe	**566,269**
NET POWER, kWe	**7,850**
CONSUMABLES	
As-Received Coal Feed, lb/hr	1,121,793
Natural Gas Feed, lbs/hr	424,461
Thermal Input, kWt	6,646,752
Condenser Duty, MMBtu/hr	845
Raw Water Withdrawal, gpm	25,608
Raw Water Consumption, gpm	23,531

8.4. Carbon Balance

Table 8-2 shows the carbon balance for the Case 3FT GTI HMB gasifier-based FT CNTL plant. Carbon emission is based on the total stack emissions from the process heaters, cogen and SRU/TGTU. CO_2 is recovered from the AGR and the PC amine CO2 recovery units with the balance of carbons going to the FT products. The waste liquid effluent and ash byproducts contain a small amount of carbon.

Table 8-2
Case 3FT Carbon Balance

OVERALL CARBON BALANCE		
Case 3FT		
Case Description	55% Coal / 45% NG Indirect External Series Reforming	
	lbs/hr	**%**
Carbon IN:		
Coal, NG & Flux	1,064,320	99.94
Cogen Combustion Air	601	0.06
Total Carbon In	**1,064,920**	**100.00**
Carbon OUT:		
Process Htr Stack	39,133	3.67
Cogen Stack	26,030	2.44
SRU/TGTU Stack	30,796	2.89
Recovered C from AGR	321,657	30.20
Export FG	0	0.00
AGR Purge H2O	111	0.01
FT Waste H2O	7,719	0.72
Slag & Ash	3,893	0.37
Naphtha + Diesel Products	488,215	45.85
PC Amine CO2 Recovery	147,503	13.85
Convergence Error	-135	-0.01
Total Carbon Out	**1,064,920**	**100.00**
Carbon Emission:		
Total Carbon Emissions	103,788	9.01

8.5. Water balance

Case 3FT water makeup and consumptions are shown in Table 8-3. The scrubber water demand is based on a maximum chlorides concentration in the scrubber purge water of 1,000 ppmw.

Table 8-3
Case 3FT Water Balance

Water Balance (GPM)	**Case 3FT 55% Coal / 45% NG Co-Feed Series Indirect Reforming**				
Water Usage by Area	Water Demand	Internal Recycle	Net Water Demand	Process Water Discharge	Raw Water Withdrawal
(GPM)					
Slag Handling	454		454	-	
SG Scrubber Makeup	5,878		5,878	(6,478)	
SWS	3,415	(3,415)	-		
SG Cooling Cond		(3,412)	(3,412)		
AGR/SRU/TGU Steam Cond		(3,963)	(3,963)	(208)	
FT Block			-	(1,337)	
BFW	17,030	(5,228)	11,802		
Blowdowns			-	(631)	
Fuel Gas Saturator	-		-		
Cooling Tower	12,724		12,724	(1,725)	
Potable Water + Contingency	48		48	8	
Total Water Usage	39,548	(16,017)	23,531	(10,372)	
Raw Water Treating					
RO/Demin System			(8,344)	(1,325)	9,621
Makeup Cooling Water Treating			(15,187)	(799)	15,987
Total Treated Water			(23,531)	(2,124)	25,608
Total	39,548	(16,017)	-	(12,496)	25,608

8.6. Capital Cost

8.6.1. Total Plant Cost

Table 8-4 shows the total plant cost (TPC) summary of GTI HMB Case 3FT. Account 4 includes the cost of the gasifier, steam methane reformer, natural gas compressors, syngas cooling and ASU.

Table 8-4
Case 3FT Total Plant Cost Summary

Total Plant Cost (June 2011)		GTI HMB Case 3FT
Acct. No.	**Item/Description**	**$MM**
1	COAL & SORBENT HANDLING	$ 73.71
2	COAL & SORBENT PREP & FEED	$ 385.31
3	FEEDWATER & MISC BOP SYSTEMS	$ 64.24
4	GASIFIER & ACCESSORIES	$ 2,211.46
5A	GAS CLEANUP & PIPING	$ 1,126.21
5AA	FT SYNTHESIS AND PRODUCT UPGRADE	$ 1,029.49
5B.2	CO2 Compression & Drying	$ 72.42
6	COMBUSTION TURBINE/ACCESSORIES	$ 151.30
7	HRSG, PC AMINE UNIT, DUCTING & STACK	$ 140.59
8	STEAM TURBINE GENERATOR	$ 79.13
9	COOLING WATER SYSTEM	$ 68.95
10	ASH/SPENT SORBENT HANDLING SYS	$ 74.54
11	ACCESSORY ELECTRIC PLANT	$ 141.45
12	INSTRUMENTATION & CONTROL	$ 36.79
13	IMPROVEMENTS TO SITE	$ 50.53
14	BUILDINGS & STRUCTURES	$ 35.84
		$ 5,741.97

8.7.Operating Costs

Table 8-5 summarizes the operating costs for GTI HMB Case 3FT.

Table 8-5
Case 3FT Operating Cost Breakdown

OPERATING COSTS, 2011 $MM/yr	GTI HMB Case 3FT
FIXED OPERATING COSTS	
Annual Operting Labor Cost	$28.9
Maintainence Labor Cost	$51.6
Administration & Support Labor	$20.1
Property Taxes and Insurance	$114.8
TOTAL FIXED OPERATING COSTS	$215.5
VRIABLE OPERATING COSTS (@90% CF)	
NON-FUEL VARIABLE OPERATING COSTS	
Maintenance Material Cost	$97.4
Water	$10.1
Chemicals	
MU & WT Chemicals	$9.7
Other Chemicals & Catalysts	$16.2
Waste Disposal	$11.2
Power Credits	($3.6)
TOTAL NON_FUEL VARIABLE OPERATING COSTS	$141.0
FUEL (@90% CF)	
Coal	$303.4
Natural Gas	$428.6
TOTAL VARIABLE OPERATING COSTS	$873.0

8.8.Cost of Electricity

Table 8-6 shows a summary of the power output, CAPEX, OPEX, COP and cost of CO_2 capture for Case 3FT.

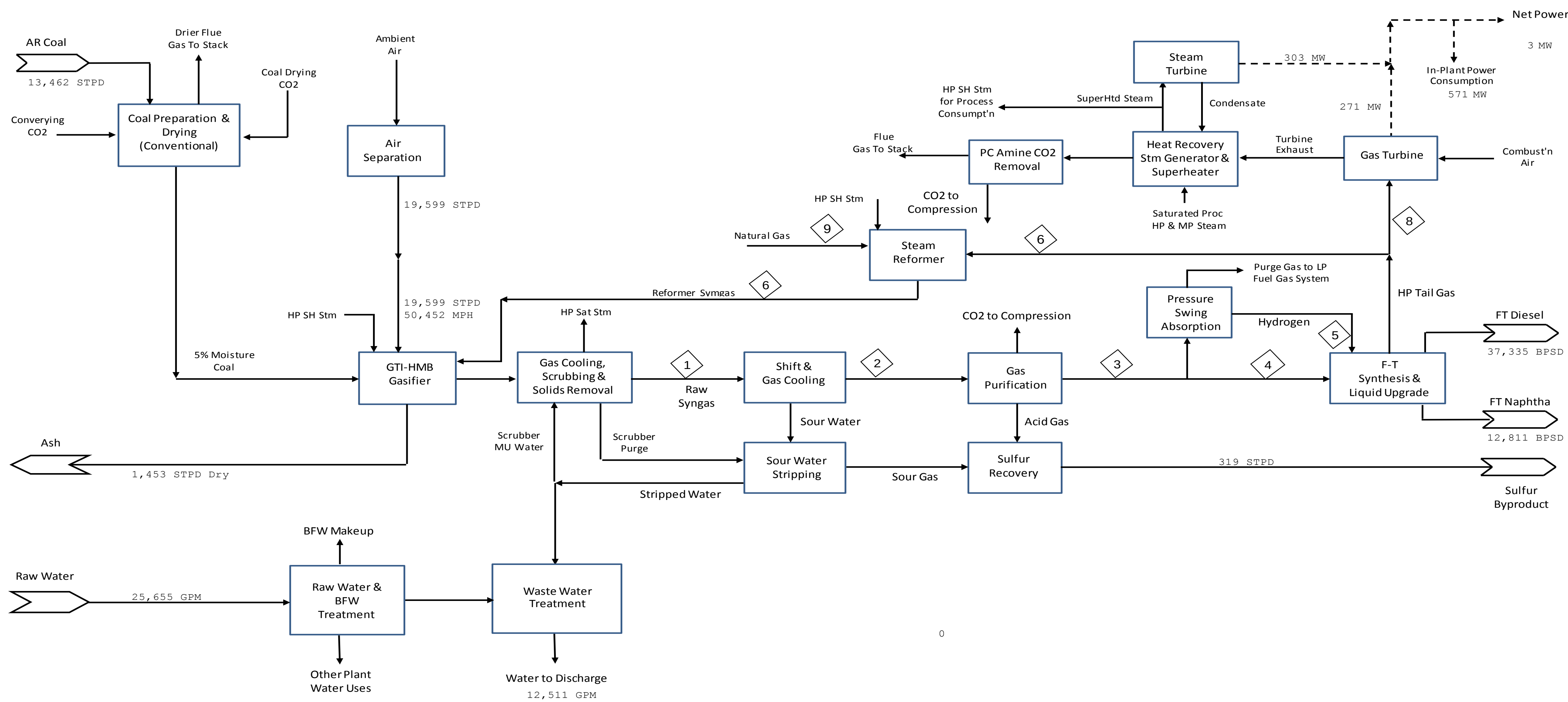
Table 8-6
Case 3FT Plant Performance and Economic Summary

	GTI HMB Case 3FT
CAPEX, $MM	
Total Installed Cost (TIC)	$4,044
Total Plant Cost (TPC)	$5,742
Total Overnight Cost (TOC)	$7,139
OPEX, $MM/yr (90% Capacity Factor Basis)	
Fixed Operating Cost (OC_{Fix})	$215
Variable Operating Cost Less Fuel (OC_{VAR})	$141
Fuel Cost (OC_{Fuel})	$732
Power Production, Mwe	
Gas Turbine	271
Steam Turbine	303
Auxiliary Power Consumption	566
Net Power Output	8
Power Generated, MWh/yr (MWH)	68,767
COP FT Diesel, excl CO2 TS&M, $/bbl FT diesel	174.3
COP FT Diesel, incl CO2 TS&M, $/bbl FT diesel	183.7
COP FT EPD, excl CO2 TS&M, $/bbl EPD	169.0
COP FT EPD, incl CO2 TS&M, $/bbl EPD	178.4
COP FT ECO, excl CO2 TS&M, $/bbl ECO	135.2
COP FT ECO, incl CO2 TS&M, $/bbl ECO	144.6
COP FT Naphtha, excl CO2 TS&M, $/bbl FT Naphtha	121.3
COP FT Naphtha, incl CO2 TS&M, $/bbl FT Naphtha	130.7

8.9.BFD & stream data

A block flow diagram with stream data for the GTI HMB Gasifier Case 3FT is shown below

Figure 8-3
BFD for GTI HMB Gasifier Case 3FT



Steam No.	1	2	3	4	5	6	8	9
Description	Raw Syngas	Shifted Syngas	Clean Syngas	Syngas to FT	H2 to Upgrade	Tail Gas to Upgrade	HP Tail Gas to GT	Natural Gas
Temperature,F	343	100	95	95	95	95	95	95
Pressure, psia	450	435	415	413	393	340	340	472
Flow, lb-mole/hr								
CO	81,193	81,193	81,193	76,774	0	13,825	7,835	0
H2	120,424	120,424	120,424	114,532	3,932	19,833	9,633	0
CO2	38,998	38,998	6,532	6,177	0	11,055	4,892	268
CH4	6	6	6	5	0	1,305	559	24,977
C2H6	0	0	0	0	0	144	62	858
C3H8	0	0	0	0	0	284	122	188
C4 +	0	0	0	0	0	321	138	107
Ar + N2	10,510	10,510	10,510	9,945	40	6,951	3,209	429
H2S	761	761	0	0	0	0	0	0
H2O	95,235	550	0	0	0	21	9	0
Total, lb-mole/hr	347,195	252,510	218,664	207,433	3,972	53,738	26,458	26,828
Total, lb/hr	6,342,199	4,636,397	3,167,599	2,996,737	9,040	1,213,077	590,497	464,869

Rev No.	Date	Revision	Engineer

Nexant, Inc.
San Francisco, California
DOE/NETL Advanced Gasification Technologies Program
GTI Hybrid Molten Bed Gasifier for Production of High Hydrogen Syngas Project
Overall Block Flow Diagram
Case 3FT
HMB Gasifier F-T CNTL Plant
56% Coal / 44% Natural Gas Co-Feed with CO2 Capture (Series Direct Reforming)

Job No.	Drawing No.	Revision No.
A02220	BFD - 003	0

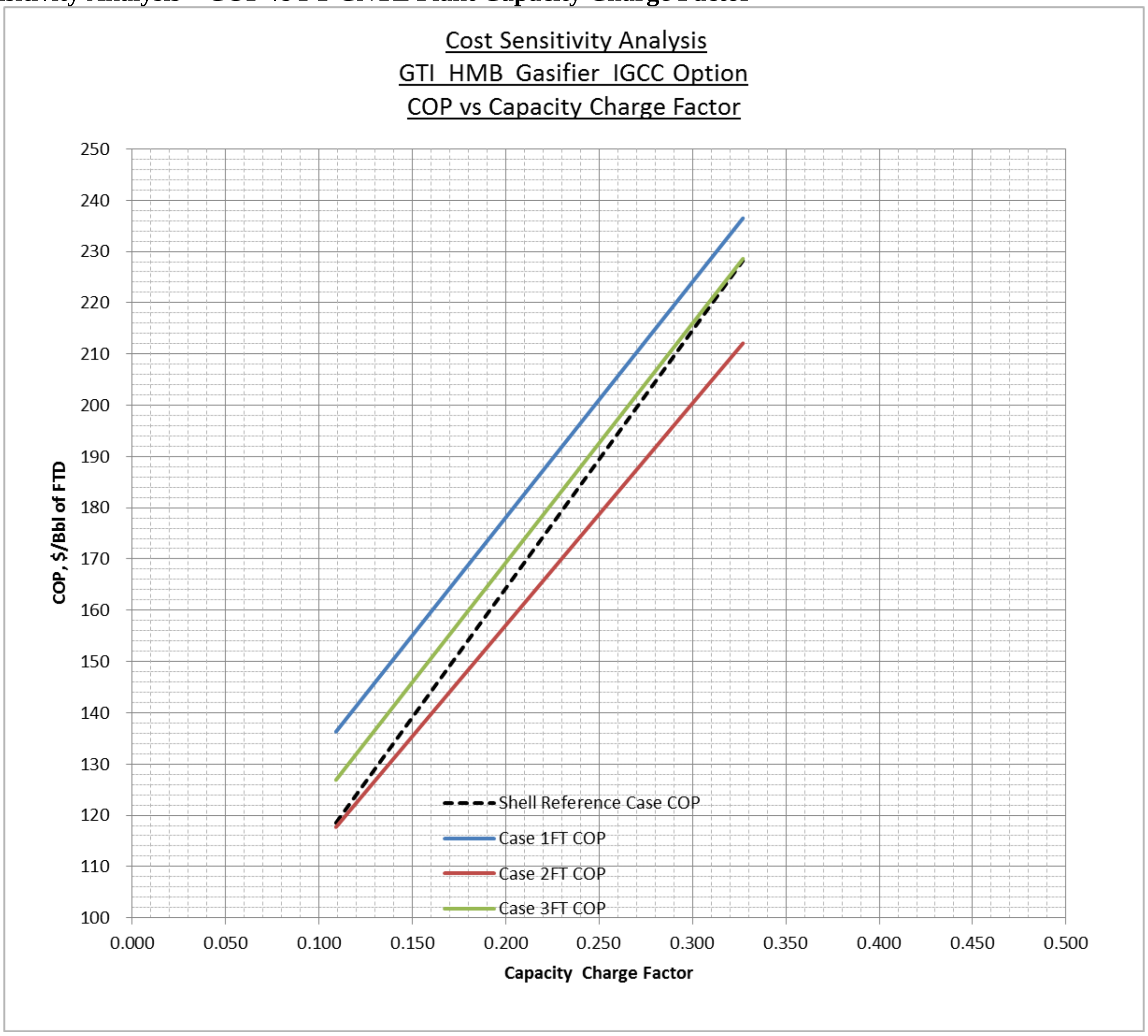
9. SENSITIVITY ANALYSIS

Sensitivity analysis was carried out to determine the effects of various parameters on the overall FT CNTL COP. The parameters investigated here include: capacity charge factor and feedstock prices.

9.1. Capacity Charge Factor

The baseline FT CNTL plant capacity charge factor used in this study is 0.218 which corresponds to the commercial fuels project finance structure with no government loan guarantees. Figure 9-2 shows the variation of FT CNTL COP with plant capacity charge factor as it varies from -50% to +50%. If government loan guarantees are assumed, the capacity charge factor is reduced to 0.17 which will reduce the COP by about 12%.

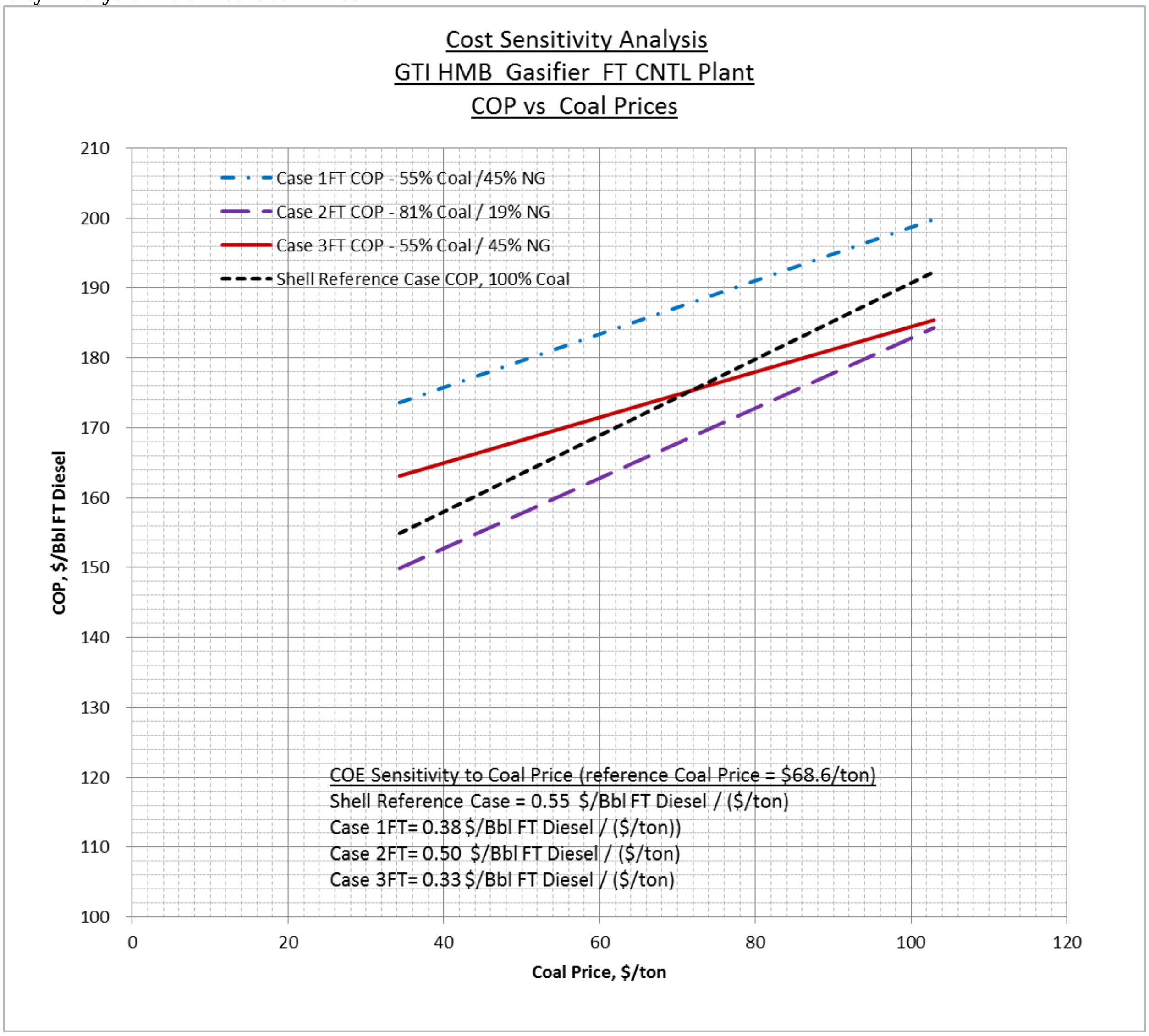
Figure 9-1
Sensitivity Analysis – COP vs FT CNTL Plant Capacity Charge Factor



9.2.Coal Price

The baseline FT CNTL plant coal price used in this study is $68.6/ton. Figure 9-3 shows the variation of FT CNTL COP with Illinois No. 6 coal price as it varies from -50% to +50% (~$34/Ton to ~$103/Ton). As expected, the COP improvement with HMB gasifier under a coal/NG co-firing situation is less compared to the 100% coal and the 81% coal cases. For example, for a 25% reduction in coal price, the improvement in COP is ~ 3% for 55% coal co-fired cases as compared to ~5% for 100% or high coal feed cases.

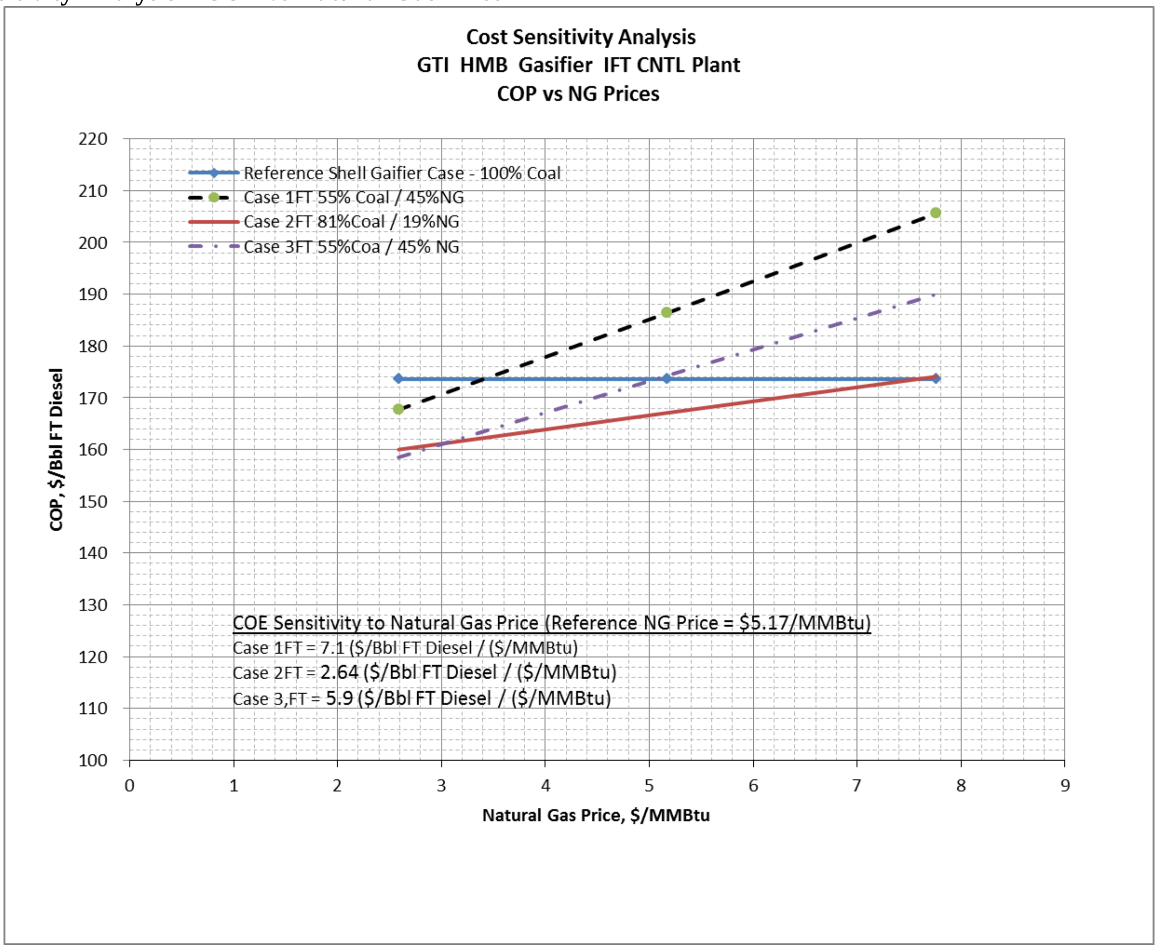
Figure 9-2
Sensitivity Analysis – COP vs Coal Price



9.3.Natural Gas Price

The baseline FT CNTL plant natural gas price used in this study is $5.34/ 1000 ft3 ($5.17/MMBtu). Figure 9-4 shows the variation of FT CNTL COP with natural gas price as it varies from -50% to +50% (~$2.6/MMBtu to ~$7.8/MMBtu). As shown, at the current spot gas market price of $2.5/MMBtu, an 11% COP reduction can be obtained, in comparison with the Reference FT CNTL case. Natural gas price impact on the COP is less for Case 2FT since the feed mix contains only 19% natural gas. Only 5% COP reduction is obtained if the natural gas price drops to $2.5/MMBtu.

Figure 9-3
Sensitivity Analysis – COP vs Natural Gas Price



10. CONCLUSIONS AND RECOMMENDATIONS

The objective of this techno-economic analysis is to assess the cost and performance of an FT CNTL plant with CO_2 capture that utilizes GTI's hybrid molten bed (HMB) gasification process to gasify Midwestern Illinois No. 6 coal. The GTI HMB gasifier is a dual coal-natural gas fueled molten bed gasification process. By varying coal and natural gas feed rates, and steam to natural gas ratio to the gasifier, the syngas H_2/CO ratio can be enhanced in the HMB gasifier to improve the overall FT CNTL plant efficiency and reduce the plant cost.

Three GTI HMB Gasifier cases with variations in feed mix, gasifier configuration and heat integration schemes are analyzed for the FT CNTL plant with CO_2 capture option. These cases are evaluated against a reference Shell SCGP gasifier based FT CNTL plant with CO_2 capture. Schematic depictions of these cases are included in the simplified block flow diagrams in figures 4-1 to 4-4 in section 4. The four cases studied are:

- Reference Case - Shell SCGP Gasifier FT CNTL Plant with 100% Illinois No. 6 Coal Feed and CO2 Capture
- Case 1FT- GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture
- Case 2FT- GTI HMB Gasifier FT CNTL Plant with 81% Coal / 19% NG Feed and CO_2 Capture (Parallel Indirect Reforming)
- Case 3FT - GTI HMB Gasifier FT CNTL Plant with 55% Coal / 45% NG Feed and CO_2 Capture (Series Indirect Reforming)

The four cases are evaluated and compared based on their overall merits in terms of their cost of production (COP).

10.1. Plant Cost of Production Results

The cost of production results of the study analysis are summarized in Table 10-1for each of the techno-economic analysis cases.

Table 10-1
FT CNTL Plant Cost and Production Summary

Case	Reference Shell Gasifier FT CTL	Case 1FT Direct Reforming FT CNTL	Case 2FT Parallel Indirect Reforming FT CNTL	Case 3FT Seriesl Indirect Reforming FT CNTL
2011 Capital Cost, $MM				
Total Plant Cost, $MM	6,543	5,702	5,571	5,742
Total Overnight Cost, $MM	8,078	7,116	6,920	7,139
2011 Operating Cost, $MM/yr				
Fixed Operating Costs	240	214	210	215
Variable Operating Costs @ 90% CF	186	147	152	141
Fuel Costs @ 90% CF, Coal @$68.6/ton	518.3	349.9	466.5	303.4
NG @ $5.17/MMBtu	0.0	506.1	191.7	428.6
Total Fuel Cost	518.3	856.0	658.2	732.0
COP FT Diesel, excl CO2 TS&M, $/bbl FT diesel	174	187	167	174
COP FT Naphtha, excl CO2 TS&M, $/bbl FT Naphtha	121	130	116	121
COP FT ECO, excl CO2 TS&M, $/bbl ECO	135	145	130	135
COP FT EPD, excl CO2 TS&M, $/bbl EPD	168	181	162	169

Of the four FT CNTL cases analyzed, Case 2FT is $7/Bbl FT diesel or 4% lower in COP relative to the reference Shell gasifier case. This case is configured with GTI HMB gasifiers and parallel indirect natural gas steam reforming and requires 81% coal and 19% natural gas as feed to produce 49.955 BPD of FT liquid fuels. Section 10-2 will compare and identify the key parameters affecting the COP based on the techno-economic performance for the four FT CNTL cases. The comparison will identify the key reasons for Case 2FT having the lowest COP.

10.2. Plant Techno-economic Results

The plant techno-economic results of the study analysis are summarized in Table 10-2 for each of the techno-economic analysis cases.

Table 10-2
FT CNTL Plant Techno-Economic Performance Summary

Case	Reference Shell Gasifier FT CTL	Case 1FT Direct Reforming FT CNTL	Case 2FT Parallel Indirect Reforming FT CNTL	Case 3FT Seriesl Indirect Reforming FT CNTL
FT Liquid Fuels Products				
FT Diesel, BPD	38,053	36,611	37,193	37,335
FT Naphtha, BPD	13,057	12,562	12,762	12,811
Total FT Liquid Fuels, BPD	51,110	49,173	49,955	50,146
Feed Mix (HHV)				
Coal, MMBtu/hr	22,360	15,094	20,126	13,087
Natural Gas, MMBtu/hr	0	12,406	4,699	10,506
Total, MMBtu/hr	22,360	27,500	24,825	23,593
% Coal	100	55	81	55
% NG	0	45	19	45
Coal, TPD As Received	23,000	15,527	20,702	13,462
Oxygen Feed, TPD (100% O2 Basis)	18,508	27,465	17,019	18,404
FT Feed Gas				
H2/CO, mol/mol	1.47	1.49	1.48	1.49
CO2 to Sequestration, TPD	31,755	27,963	27,592	20,643
Power Production, Mwe				
Gas Turbine	260	190	414	271
Steam Turbine	407	579	161	303
Auxiliary Power Consumption	664	763	574	566
Net Power Output	3	6	1	8
Raw Water Withdrawal, gpm	47,729	32,072	34,280	25,608
No. of Gasifiers (including spares)	9	22	11	18
No. of Reformer	0	0	12	14
No. of ATR	4	0	0	0

The relative oxygen feed requirement is a very good indication of the gasification efficiency and gasification system cost when comparing gasification processes. Case 2FT requires the least amount of oxygen feed among the four cases. This is the result of gasifying coal only in the gasifier and reforming natural gas with steam external to the gasifier. In other cases, oxygen is used to gasify coal and also gasify and heat the co-feeds to the gasifier. Hence, more oxygen is required for the other cases. The following compares Case 2FT which has the lowest oxygen feed to the other three FT CNTL cases:

- The reference Shell case has higher coal feed rate than Case 2FT (100% coal feed (23,000 TPD) vs 81% coal feed mix (20,702 TPD) for Case 2FT). More oxygen is required.

- Although Case 1FT uses less coal (55% coal mix, 15,527 TPD vs. 20,702 TPD), it also uses natural gas and steam as co-feed and the resulting syngas must also be heated to the gasification temperature of 2,600°F. Hence, more oxygen is required to gasify and heat the coal, natural gas and steam co-feeds to the gasifier for Case 1FT.

- Case 3FT uses the least coal (55% coal mix, 13,462 TPD) and also has the advantage of recycling the heat from the reformer syngas. It requires more oxygen than Case 2FT because of the need to also heat the recycled reformer syngas to 2,600°F. This is reflected in this case by having the second lowest oxygen requirement.

The oxygen requirement is also an indication of the amount of gasification syngas generated which impacts the size and cost of the gasification trains. The gasification train is consisted of the gasifiers, ASU, coal handling and conveying, natural gas compression, steam reformer, syngas heat recovery and syngas cleaning.

The number of GTI HMB gasifiers determines the size of the gasification train. The number was estimated based on the gasifier syngas rate and the residence time of 4 seconds per gasifier. It can be seen that Case 1FT has the highest number of HMB gasifiers (22) and case 2FT has the lowest number of HMB gasifiers (11). The impact of the gasification train on the TPC for the gasification system is shown in Table 10-3.

The lower number of gasifiers for Case 2FT corresponds to the lower cost of the HMB gasifiers and the gasification system TPC. It can be seen in Table 1-3 that the lower cost of the gasifiers for Case 2FT along with the lower cost of the ASU, natural gas compression and syngas heat recovery offset the additional cost of the reformers.

Table 10-3
Gasifier & Accessories Account Details

TOTAL PLANT COST, 2011 $MM	Reference Shell Gasifier FT CTL	Case 1FT Direct Reforming FT CNTL	Case 2FT Parallel Indirect Reforming FT CNTL	Case 3FT Seriesl Indirect Reforming FT CNTL
GASIFIER & ACCESSORIES				
Gasifier, Quench Column, Filters & Cyclones	1,671.4	896.3	488.2	798.9
Steam Reformer	-	-	472.3	551.0
Natural Gas Compression	-	17.4	8.0	9.3
Syngas Heat Recovery	-	106.7	30.5	61.3
ASU/Oxidant Compression	673.4	789.0	619.1	653.9
LT Heat Recovery & FG Saturation	85.9	69.0	44.9	60.3
Flare Stack System	6.6	5.4	6.2	5.0
Gasification Foundations	93.8	77.1	89.0	71.8
Total	2,531.1	1,960.9	1,758.2	2,211.5
No. of Gasifiers (including spares)	9	22	11	18
No. of Reformer	-	-	12	14
No. of ATR	4	-	-	-

10.3. Conclusion

In conclusion, Case 2FT with parallel indirect reforming is recommended for further study and development because of its lower COP. Areas where further cost reductions are possible are in the further development of the gasifier and gas to gas steam methane reformer design

Appendix B Acronyms and Abbreviations

°F	Degree Fahrenheit
AGR	Acid Gas Removal
AOI	Area of Interest
AR	Aerojet Rocketdyne
Ar	Argon
ASU	Air Separation Unit
ATR	Autothermal reformer
Bbl	Barrels
BEC	Bare Erected Cost
BFD	Block Flow Diagram
BFW	Boiler Feed Water
B/L	Battery Limit
BOP	Balance of Plant
BPD	Barrels per day
Btu	British Thermal Unit
CAPEX	Capital Expenditure
CCF	Capital Charge Factor
CF	Capacity Factor
CH_4	Methane
CGE	Cold Gas Efficiency
Circ	Circulating
CNTL	Coal/Natural Gas to Liquid
CO	Carbon Monoxide
CO_2	Carbon Dioxide
COP	Cost of Production
COS	Carbonyl Sulfide
CT	Combustion Turbine
CTL	Coal to Liquid
CTG	Combustion Turbine Generator
CW	Cooling Water
DBT	Dry Bulb Temperature
DI	De-ionized
DOE	U.S. Department of Energy
DSP	Dry Solids Pump
ECO	Equivalent Crude Oil
EPC	Engineering, Procurement and Construction
EPD	Equivalent Petroleum Diesel

EPRI	Electric Power Research Institute
°F	Degrees Fahrenheit
FG	Fuel Gas
FO	Fuel Oil
FOA	Funding Opportunity Announcement
ft	feet
FT	Fischer-Tropsch
GE	General Electric
GT	Gas Turbine
GTG	Gas Turbine Generator
GTI	Gas Technology Institute
h	Hour
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen Sulfide
Hg	Mercury
HGCU	Hot Gas Clean Up
HHV	Higher Heating Value
HMB	Hybrid Molten Bed
HP	High Pressure
HRSG	Heat Recovery Steam Generator
HTGC	High Temperature Gas Cooling
In Hg	Inches of mercury
I & C	Instrumentation & Control
IGCC	Integrated Gasification Combined Cycle
IOU	Investor Owned Utility
IP	Intermediate Pressure
ISO	International Organization for Standards
kg	kilogram
kWe	Kilowatt electric
kWh	kilowatt hour
lb	Pound Mass
LH	Lock Hopper
LNB	Low NO_x Burner
LP	Low Pressure
LTGC	Low Temperature Gas Cooling
max	Maximum
ME	Major Equipment
MEC	Major Equipment Cost
min	Minimum

Misc	Miscellaneous
MM	million
MNQC	Multi-Nozzle Quiet Combustor
MP	Medium Pressure
MU	Makeup
MW	Megawatt
MWe	Megawatt electric
MWh	megawatt hour
N_2	Nitrogen
NETL	National Energy Technology Laboratory
NG	Natural Gas
NGB	Non-GTI Block
NOx	Oxides of Nitrogen
NPE	Net Plant Efficiency
NSPS	New Source Performance Standards
O&M	Operating and Maintenance
O_2	Oxygen
OPEX	Operating Expenditure
OSBL	Outside Battery Limit
PC	Pulverized Coal
PFD	Process Flow Diagram
PM	Particulate Matter
POTW	Public Owned Treatment Works
ppmv	Parts per Million by Volume
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
SC	Supercritical
scf or SCF	Standard Cubic Feet
scfh or SCFH	Standard Cubic Feet per Hour
scfm or SCFM	Standard Cubic Feet per Minute
SCGP	Shell Coal Gasification Process
SCM	Submerged Combustion Melting
SO_2	Sulfur Dioxide
SOPO	Statement of Project Objectives
ST	Steam Turbine

STG	Steam Turbine Generator
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	Short Tons per Day
TS&M	Transportation, Storage & Maintenance
US, USA	United States of America
vol%	Percentage by Volume
WBT	Wet Bulb Temperature
WGS	Water Gas Shift
WT	Waste Treatment
WTA	Licensed Coal Drying Process
wt%	Percentage by Weight