Final Scientific/Technical Report

Project Title:	Advanced Acid Gas Separation Technology for Clean Power and Syngas Applications
Reporting Period:	October 1, 2013 to June 30, 2015
Report Submission Date:	17 September 2015
Federal Agency/Org. Element:	DOE/National Energy Technology Laboratory
DOE Award Number:	DE–FE0013363
Project Period:	October 1, 2013 to June 30, 2015
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09-17-2015

Dr. Fabrice Amy Date

DISCLAIMER

Advanced Acid Gas Separation Technology for Clean Power and Syngas Applications

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Abstract

Air Products has developed an acid gas removal technology based on adsorption (Sour PSA) that favorably compares with incumbent AGR technologies. During this DOE-sponsored study, Air Products has been able to increase the Sour PSA technology readiness level by successfully operating a two-bed test system on coal-derived sour syngas at the NCCC, validating the lifetime and performance of the adsorbent material. Both proprietary simulation and data obtained during the testing at NCCC were used to further refine the estimate of the performance of the Sour PSA technology when expanded to a commercial scale. In-house experiments on sweet syngas combined with simulation work allowed Air Products to develop new PSA cycles that allowed for further reduction in capital expenditure. Finally our techno economic analysis of the use the Sour PSA technology for both IGCC and coal-to-methanol applications suggests significant improvement of the unit cost of electricity and methanol compared to incumbent AGR technologies.

Appendix

Executive Summary

Gasification is a promising alternative to traditional coal-fired combustion that can be adapted to CO_2 capture while supplying synthesis gas (syngas) for hydrogen, power, or chemical products. A key challenge for coal gasification is to reduce its cost. Downstream processing of syngas for CO_2 capture requires separation of the crude stream into the desired products (H_2/CO), a sulfur stream (primarily H_2S), and sequestration-ready CO_2. The process most commonly employed for this separation is acid gas removal (AGR) based on absorption in a physical solvent. Air Products has developed a proprietary alternative that consists of two process blocks: 1) sour pressure swing adsorption (PSA) that separates CO_2 and H_2S from the desired products, and 2) a tail gas disposition block that separates the sulfur-containing compounds and purifies the CO_2 to a sequestration-grade product. Sour PSA is the key enabler of the technology, but only limited testing of the adsorbent technology had been performed on high-hydrogen syngas streams.

In this project, Air Products built and operated a two-bed mobile PSA unit at the National Carbon Capture Center (NCCC) facility in Wilsonville, AL. A slipstream of authentic, high-hydrogen syngas derived from low-rank coal was used as the feedstock. Testing was conducted over 26 days during the October-November 2014 gasification run at NCCC, totaling more than 2,000 PSA cycles. By utilizing real-world, high-hydrogen syngas, information necessary to understand the utility of the system for IGCC application and methanol production was made available. The PSA rejected nearly all of the H_2S (>98.5%) and more than 90% of the CO_2 in the feed gas to the depress/purge effluent gas. Ethane and carbonyl sulfide (COS) also preferentially partition to tail gas effluent gas, while N_2, CO, and CH_4 tend to partition to both the product gas and waste gas. Product gas purity for the sour syngas cases was maintained at <6 ppm H_2S even though the feed level was ~300 ppm. As expected, attainment of higher-purity product gas was achieved with reduced feed loading. One of the key goals of this test was to determine if the PSA performance was stable after the adsorbent was exposed to sour syngas. As a check, the PSA was operated at the end of the campaign with the same process setpoints used at the beginning of the campaign. The PSA system yielded essentially the same product composition, indicating that the adsorbent characteristics remained stable throughout the test period.

Postmortem analysis of the adsorbent material revealed that the levels of detectable species on the exposed and fresh samples were similar, indicating no accumulation on the exposed adsorbent. In addition, the adsorbent surface area and pore volume were determined by conducting N_2 adsorption measurements. The surface area of the exposed samples was similar to that of the fresh sample, and the pore volume of the used samples appeared to increase only slightly. There was therefore no indication of adsorbent degradation due to extended sour syngas exposure. When combined with prior knowledge, these experimental results allowed Air Products to estimate that in commercial applications, the lifetime of the adsorbent material would be at least 10 years.

In addition, Air Products' existing multi-bed PSA system (internally referred to as the process development unit, or PDU) was used to experimentally validate new PSA process cycles for generating syngas suitable for methanol synthesis or hydrogen production for power generation. Since that unit could not accommodate sulfur species, it was operated with sweet syngas at flows and pressures relevant to industrial operation. The motivation for developing these new PSA process cycles was to reduce equipment cost by directly increasing product (H_2, H_2+CO) recovery from the PSA unit. Indeed, increasing PSA direct recovery alleviates the need for equipment to recover product slipping into the tail gas. A PSA cycle designed to maximize the recovery of H_2 as well as CO was tested for methanol production applications. The cycle was unique since it included a CO_2 rinse step after the feed step. The rinse cycle yielded H_2 recoveries >95%, while CO recovery generally fell in the 80-90% range.

For IGCC applications, H_2 recovery was maximized by adding a N_2 rinse step after the feed step. The recovery of H_2 was experimentally demonstrated to be very high (97 %) with this improved PSA cycle. The major components of the product stream consisted of ~85% H_2 and 15% N_2. Further dilution of this product stream with N_2 and/or steam would result in a suitable fuel for a gas turbine in power generation applications. However, N_2 slip into the tail gas would result in the CO_2 stream not meeting the required criteria for sequestration. It was therefore concluded that for IGCC applications, a classical PSA cycle (i.e., not including the N_2 rinse step) would be used, resulting in H_2 recovery in the mid-90% range. Air Products' proprietary process simulator (SIMPAC) was used to predict the performance of these PDU tests. The agreement between experiments and simulation was very good, giving confidence in the estimated performance of the process when scaled up to commercial levels.

Techno-economic assessments (TEAs) of the performance of the Sour PSA technology were developed for both IGCC and methanol production following guidelines published by NETL/DOE. In both cases, a baseline or reference case using conventional technology for acid gas removal was first established. Then, the TEA for a plant utilizing the Sour PSA technology was established using the same assumptions, allowing for direct comparison and estimate of the financial benefits of the new technology. For all four cases, the financial analysis was performed via cash flow analysis. Both IGCC and methanol production applications show the same trend: significant cost reduction in the acid gas removal scope allows for a reduction in the product cost. Indeed, the use of Air Products' Sour PSA technology instead of incumbent solvent-based acid gas removal technology would result in a 4.6% lower cost of electricity and a 3.7% lower cost of methanol. In addition, the cost benefit of Sour PSA technology was found to be insensitive to equipment cost, performance or the CO_2 economic environment, further validating the cost advantage of Air Products' Sour PSA technology.

Introduction

Project Objectives

Air Products has developed a potentially ground-breaking technology – Sour Pressure Swing Adsorption (PSA) – to replace the solvent-based acid gas removal (AGR) systems currently employed to separate sulfur-containing species, along with CO_2 and other impurities, from gasifier syngas streams. The Sour PSA technology is based on adsorption processes that utilize pressure swing or temperature swing regeneration methods. Sour PSA technology has already been shown to provide a significant reduction in the cost of CO_2 capture for power generation. The objective of this project was to test the performance and capability of the adsorbents in a test unit, using a Hydrogen-rich syngas generated from the gasification of coal. In addition, Air Products operated a multi-bed PSA process development unit (PDU) to evaluate the incorporation of pressure equalization steps in the PSA cycle and to improve predictions of system operation at commercial scale. The results were used to develop a techno-economic assessment evaluating the applicability of the technology to IGCC and methanol production plants. It is anticipated that replacing the conventional AGR process with Air Products' innovative downstream processes will reduce the cost of capital for CO_2 capture in a high-hydrogen syngas, while maintaining >90% CO_2 capture efficiency.

Background

Gasification is a promising alternative to traditional coal-fired combustion that can be adapted to CO_2 capture while supplying synthesis gas (syngas) for hydrogen, power, or chemical products. A key challenge for coal gasification is to reduce its cost. Downstream processing of syngas for CO_2 capture requires separation of the crude stream into the desired products (H_2/CO), a sulfur stream (primarily H_2S), and sequestration-ready CO_2. The process most commonly employed for this separation is acid gas removal (AGR) based on absorption in a physical solvent. Air Products has developed a proprietary alternative that consists of two process blocks: Sour Pressure Swing Adsorption (PSA) that separates CO_2 and H_2S from the desired products, and a tail gas disposition block that separates the sulfur-containing compounds and purifies the CO_2 to a sequestration-grade product (Figure I-1). Sour PSA is the key enabler of the technology, but only limited testing of the adsorbent technology has been performed on high-hydrogen syngas streams.

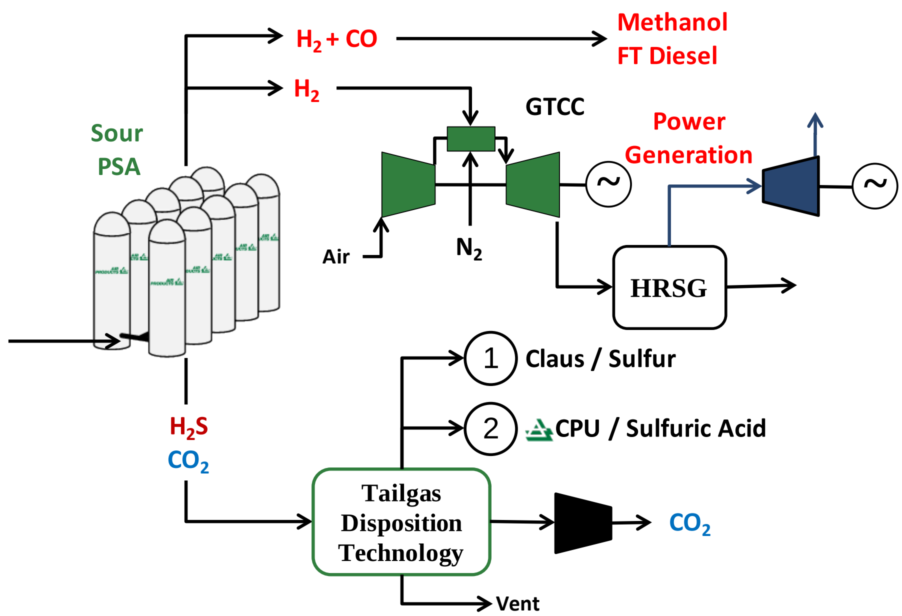


Figure I-1. The Sour PSA system.

Under DOE project DE-FE0007759, Air Products successfully demonstrated a 4% reduction in the cost of electricity (COE) from a low-rank coal-based, integrated gasification and combined cycle (IGCC) plant with CO_2 capture [1]. This represented a 14% decrease in the cost of carbon capture compared to the NETL S3B reference case [2] on a transmission, storage and monitoring (TS&M) charge basis. However, in that project, PSA operation never reached cyclic steady state due to tar accumulation in the PSA adsorbent system. In a preliminary investigation during that project, the operation of an upstream adsorption process, using either temperature swing adsorption (TSA) or a guard bed, was shown to be an effective tool for tar removal. However, due to the time and resource limitations of that project, the operation of a protective adsorption system coupled with the PSA system could not be tested. In addition, those tests were conducted only in one-week trials, a critical limitation that must be addressed prior to commercialization.

Proposed Approach

In this project, Air Products proposed to operate a two-bed mobile PSA unit, with an upstream guard bed, at the National Carbon Capture Center (NCCC) facility. A slipstream of authentic, high-hydrogen syngas based on low-rank coal would be evaluated as the feedstock. Testing was to be conducted for approximately eight weeks, thereby providing far longer adsorbent exposure data than demonstrated to date. By utilizing real-world, high-hydrogen syngas, information necessary to understand the utility of the system for methanol production would be made available.

In addition, Air Products would also operate a multi-bed PSA process development unit (PDU), located at its Trexlertown, PA headquarters, to evaluate the impact of incorporating pressure equalization steps in the process cycle. This testing would be conducted utilizing a sulfur-free, synthetic syngas, and would improve the reliability of predictions of the system's operating performance at commercial scale.

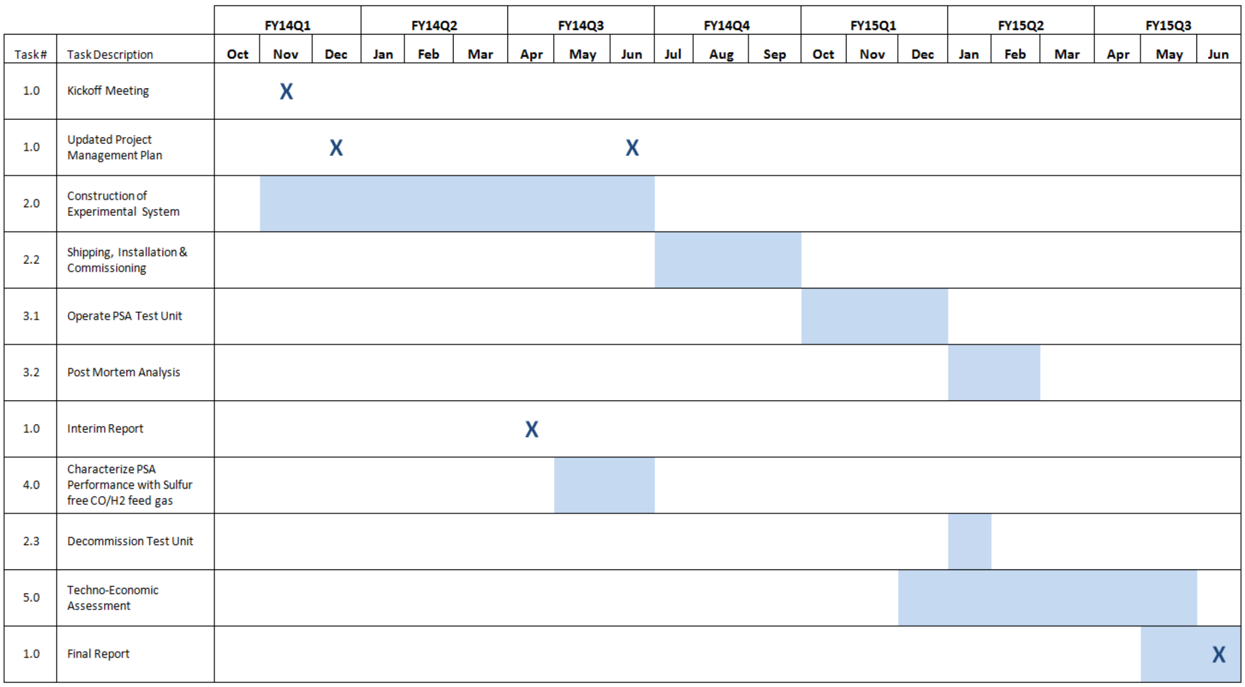
Finally, information obtained from both the two-bed PSA system operated with authentic high-hydrogen syngas and the multi-bed system operated with synthetic syngas would be combined to build a techno-economic analysis (TEA) of PSA utilization for IGCC applications and coal-to-methanol production.

Task 1: Project Management and Planning

Timeline

The project was initially scheduled for a duration of one year starting in October 2013. However, due to the impossibility of scheduling a test run at the National Carbon Capture Center before October 2014, a no-cost extension until June 2015 was granted. Table 1-1 represents the actual timeline of the project broken down by major tasks.

Table 1-1. Project timeline.



Budget and Cost Share

The overall budget for the project was $1,064,696, with an 80/20 cost share between DOE and Air Products.

Table 1-2. Project overall budget.

Air Products	$ 212,939 (20%)
DOE	$ 851,757 (80%)
Total	$ 1,064,696

The actual spending schedule for the entire duration of the project is provided in Table 1-3.

Table 1-3. Project actual budget (as of 06-30-2015) and spending schedule.

Baseline Reporting Quarter	Budget Period 1							
	FY14Q1		FY14Q2		FY14Q3		FY14Q4	
	Q1	Total	Q2	Total	Q3	Total	Q4	Total
Baseline Cost Plan								
Federal Share	112,815	112,815	162,980	275,795	209,878	485,673	51,985	537,658
Non-Federal Share	28,204	28,204	40,745	68,949	52,470	121,419	12,996	134,415
Total Planned	141,019	141,019	203,725	344,744	262,348	607,092	64,981	672,073
Actual Incurred Cost								
Federal Share	200,217	200,217	290,072	490,289	68,687	558,976	50,710	609,686
Non-Federal Share	50,055	50,055	72,519	122,574	17,172	139,746	12,678	152,424
Total Incurred Costs	250,272	250,272	362,591	612,863	85,859	698,722	63,388	762,110
Variance								
Federal Share	(87,402)	(87,402)	(127,092)	(214,494)	141,191	(73,303)	1,275	(72,028)
Non-Federal Share	(21,851)	(21,851)	(31,774)	(53,625)	35,298	(18,327)	318	(18,009)
Total Variance	(109,253)	(109,253)	(158,866)	(268,119)	176,489	(91,630)	1,593	(90,037)

Baseline Reporting Quarter	Budget Period 2					
	FY15Q1		FY15Q2		FY15Q3	
	Q1	Total	Q2	Total	Q3	Total
Baseline Cost Plan						
Federal Share	210,060	747,718	91,554	839,272	12,483	851,755
Non-Federal Share	52,515	186,930	22,888	209,818	3,121	212,939
Total Planned	262,575	934,648	114,442	1,049,090	15,604	1,064,694
Actual Incurred Cost						
Federal Share	93,290	702,976	70,756	773,732	28,941	802,673
Non-Federal Share	23,322	175,746	17,690	193,436	7,235	200,671
Total Incurred Costs	116,612	878,722	88,446	967,168	36,176	1,003,344
Variance						
Federal Share	116,770	44,742	20,798	65,540	(16,458)	49,082
Non-Federal Share	29,193	11,184	4,928	16,382	(4,114)	12,268
Total Variance	145,963	55,926	25,996	81,922	(20,572)	61,350

Task 2: Construction of the Two-Bed PSA Test Unit

The experimental Air Products (AP) unit was built and operated to 1) evaluate the performance of a simple two-bed Sour PSA unit to reject acid gases from real gasifier-produced syngas; and 2) determine the stability of the adsorbent in the PSA during the course of the syngas exposure. Air Products' approach was to take a small slip stream (~1 scfm, or <5 lb/hr) of sour syngas from the NCCC Syngas Conditioning Unit (SCU) and process it in the Sour PSA system. Before arriving at the AP unit, the syngas was filtered, shifted (underwent a water gas shift reaction: $CO+H_2O \Leftrightarrow CO_2+H_2$), and cooled to ~50^{o}F. The cooling step generated condensate, which was rejected along with significant amounts of waxy, organic tar species. Thankfully, these tasks were handled by NCCC personnel who have years of experience dealing with these operations. The dewatered sour syngas was delivered to the Air Products area at ambient temperature and ~150 psig.

Processing the sour syngas generated essentially two process streams: a PSA product stream and a PSA depress/purge gas stream. These were recombined in a waste gas header and sent back to NCCC for destruction in their thermal oxidation unit. Waste gas generated from other steps, like N_2 purging or guard bed regeneration, were handled the same way. Thus, there were no significant emissions of sour syngas from the AP unit.

A simple schematic of the Sour PSA test unit is presented in Figure 2-1. The main process equipment resided on the guard bed skid or the PSA skid. Raw syngas from NCCC first entered the guard bed skid, which contained manual and actuated valves for feed gas flow isolation and inert gas purging, as well as two parallel restrictive orifices to limit the worst-case flow of syngas to the unit. A simple knockout vessel was used to reject any liquids from the feed (by manually draining). The syngas pressure was reduced across a pressure regulator to ~125 psig and passed to the guard beds. Flow could be passed through one or both guard beds (Beds 1 and 2) by appropriate operation of manual ball valves. Selection of the valves also permitted thermal regeneration of one bed while the other was on feed. Hot nitrogen from NCCC was used to conduct the regeneration. The guard beds were traced with mineral-insulated heat tape and fiberglass insulation blankets to enable operation at a constant temperature, typically 30^{o}C. The heaters were also used during the thermal regeneration process. The purpose of the guard beds was to remove organic tar components from the sour syngas.

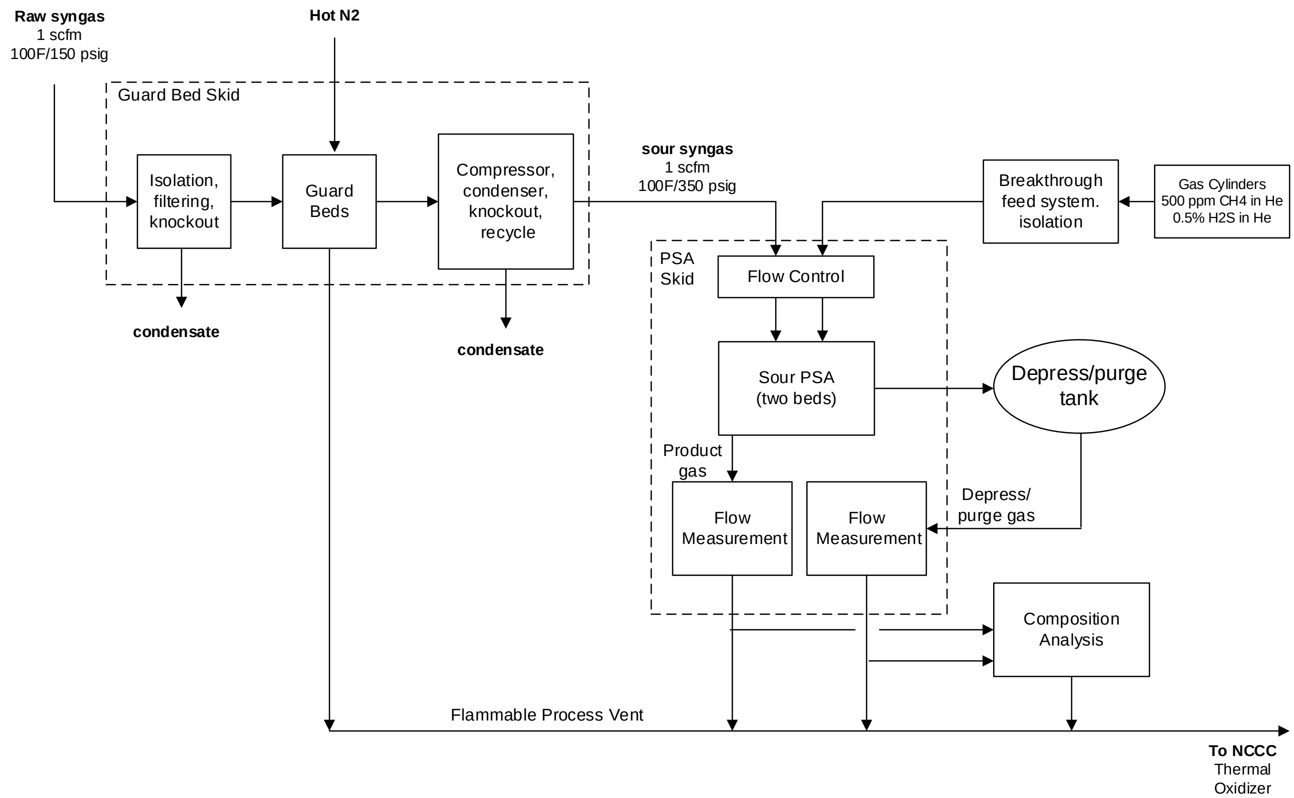


Figure 2-1. Simple schematic of the Sour PSA test unit.

After exiting the guard beds, the sour syngas passed through a set of filters (in parallel configuration) and an electrically classified diaphragm compressor for compression to ~350 psig. Water and other condensables were rejected in a manually operated, air-cooled cooler/knockout assembly. The sour syngas was then passed to the Sour PSA skid.

The inlet gas pressure was controlled with a pressure regulator to ~330 psig, and the flow was set to the desired level with a thermal mass flow controller. The Sour PSA skid included two PSA vessels containing solid adsorbent for the removal of sulfur species and CO_2. Each bed was subjected to a series of process steps including a high-pressure feed step, a countercurrent depressurization step to atmospheric pressure, a countercurrent purge step with product gas, and a countercurrent repressurization step with product gas. One bed was on the feed step while the other simultaneously underwent the regeneration steps. Execution of the cycle was handled by a process logic controller (PLC). Process variables included the duration of the various steps, the feed flow rate, the purge gas flow rate (via an electronic needle valve), the repress rate (via an electronic needle valve), and the depressurization gas flow rate (via a set of parallel electronic needle valves). Gas pressure during each step was controlled by back pressure regulators.

Gas flow rates and compositions were determined for the feed, product, and combined purge/depress effluent streams. The composition and flow of the purge/depress stream was dampened by passing them to a tank for flow averaging. Flow rates were determined by dry test meters fitted with rotary encoders for electronic signal generation. Syngas compositions were measured with an Inficon Micro-GC gas analyzer unit (Model 3000), which provided analysis every 2-3 minutes and could sample the PSA feed, product, or depress/purge gas. A lead acetate H_2S analyzer (Model 903, Galvanic Applied Sciences) was used to continuously measure ppm levels of H_2S in the PSA product gas. An Agilent 3000 Micro GC monitored the level of organic species, especially benzene, in the feed and effluent streams from the guard beds.

The pressure, temperature, and flow data were continuously logged by the PLC. Composition data were logged either by the PLC or with a dedicated PC, depending on the analyzer. Collection of flow and composition data permitted evaluation of overall and component mass balances and H_2 recovery/H_2S rejection.

The system was also used to measure breakthrough of H_2S or CH_4 in an inert helium carrier gas. In this case, feed gas was supplied from a gas cylinder, and the mass flow controller was used to deliver the gas at a fixed flow rate to one of the adsorber columns. The adsorbent was previously regenerated by purging with N_2 and pressurized to 400 psig in He. Pressure in the column was maintained at 400 psig, while the carrier gas with the desired contaminant flowed through the column. Breakthrough of the contaminant gas was monitored with an analyzer. A Horiba VIA510 infrared detector was used to detect CH_4, while either the Inficon Micro-GC or an on-line photometric H_2S analyzer (Ametek 9900) monitored the progression of H_2S. Flow continued until the concentration of the contaminant gas in the effluent stream was >60% of the feed gas composition.

A plot plan of the equipment layout is shown in Figure 2-2.

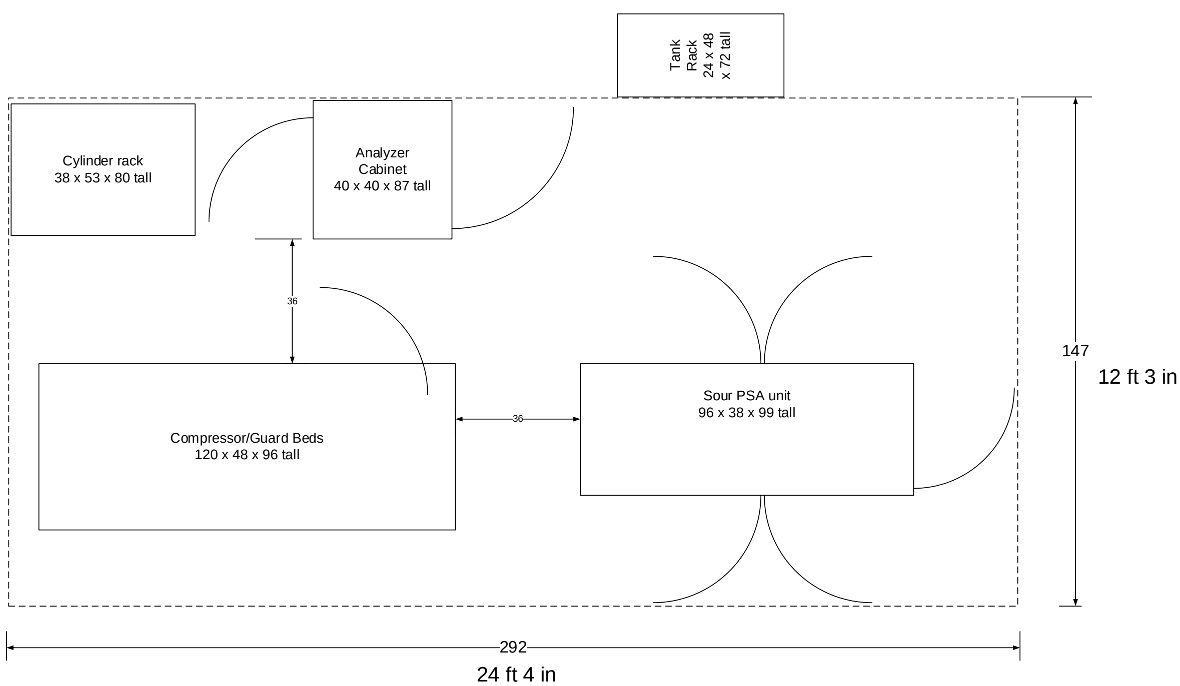


Figure 2-2. Layout of the Sour PSA unit at the NCCC site.

The equipment had to meet the Class 1 Div 2 Group B electrical classification associated with the NCCC location. The Sour PSA skid was pre-existing and contained general-purpose electrical equipment, so it was reconfigured to be continuously purged with nitrogen. A differential pressure switch was installed to de-energize the enclosure if the internal pressure dropped to < 0.15" water. The micro-GC units, also general purpose, were installed in a nitrogen-purged cabinet fitted with a commercial Type X purge system. The lead acetate tape H_2S analyzer was classified for Class 1 Div II location, as was all of the equipment installed on the guard bed skid.

The experimental test unit was designed, constructed and function tested at Air Products, then shipped to Wilsonville, AL and installed at the NCCC site. Very thorough safety evaluations were executed for this project. During the construction phase at Air Products, an internal HAZOP analysis was conducted that focused on the internal operation of the Sour PSA unit. A second HAZOP was conducted at NCCC with Air Products and NCCC personnel to review the first HAZOP and investigate specific issues associated with locating the unit at the plant. Photographs of the installation are presented in Figure 2-3.

Figure 2-3. Photos of the Sour PSA process unit.

Task 3: Operation of the Two-Bed PSA Test Unit at the NCCC

The Sour PSA system was installed at the NCCC site during summer 2014. Helium leak testing, alarm testing, and general commissioning were completed in September. Operation of the system began in October 2014.

The two PSA columns were packed with adsorbent at Air Products. Each column consisted of a 7 ft long piece of 1" OD stainless steel tube (0.065" wall). Sample and thermocouple taps were located 15, 30, 45, 60 and 72 inches from the feed end of the column. Each column was packed with 84" of Air Products' preferred sour gas adsorbent.

The guard beds were packed at the NCCC site. These vessels were 10" in diameter by ~2 ft long (10 gallon volume). Each was packed with 24 lbs of activated carbon.

The entire system was thoroughly purged with house N_2 to remove oxygen from the process lines before any flammable gases were introduced to the unit. The PSA beds were also purged to remove any adsorbed water from the adsorbent.

Initial Breakthrough Tests

Breakthrough tests were conducted with both PSA beds to evaluate the initial adsorption capacity of the sour gas adsorbent. Breakthrough runs were conducted at 400 psig/30°C with a feed gas of 2% CH_4 in He or 0.5% H_2S in He. Breakthrough with H_2S was chosen since it is a key component to be removed, while CH_4 breakthrough was selected to see if there was any difference between polar and nonpolar adsorbates. The gas feed rate was 10 slpm. The entire packed bed was used for the CH_4 runs, whereas sample gas was withdrawn from the 45" sample tap for the H_2S tests. Typical breakthrough curves are illustrated in Figure 3-1 for CH_4 and Figure 3-2 for H_2S.

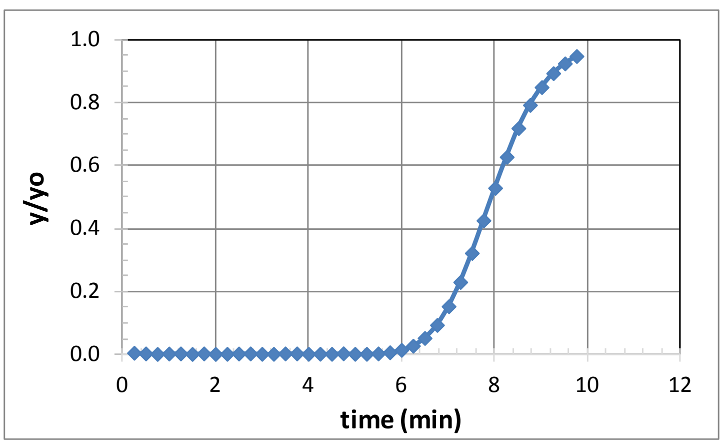


Figure 3-1. Typical breakthrough curve for 2% CH_4 in He.

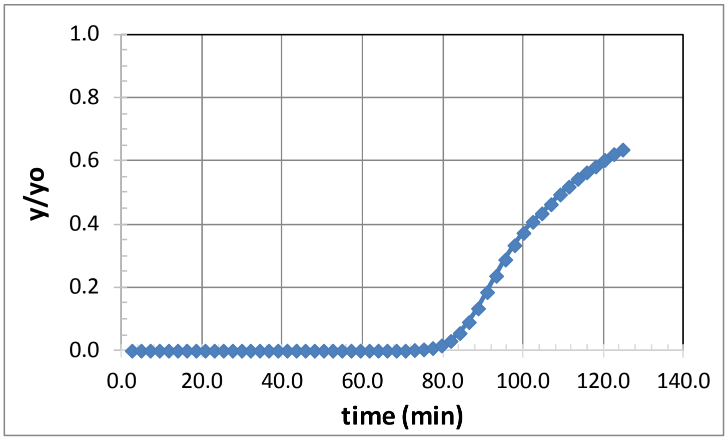


Figure 3-2. Typical breakthrough curve for 500 ppm H_2S in He.

The effective CH_4 and H_2S adsorption capacities were evaluated from the breakthrough results with the equation

$$n = \frac{t_s F y}{M_s}$$

where n is the adsorption capacity (mmole/g), t_s is the stoichiometric time where the mole fraction of adsorbable species in the column effluent is half the feed mole fraction, F is the molar feed flow rate, y is the mole fraction of adsorbed species in the feed gas, and M_s is the mass of solid adsorbent in the packed bed. The measured adsorption capacities are listed in Table 3-1. The H_2S data are plotted against the H_2S isotherm determined previously from a volumetric isotherm unit (Figure 3-3). The breakthrough capacities are in good agreement with the isotherm.

Table 3-1. Summary of initial breakthrough capacities.

Relative CH_4 Breakthrough Capacities

Run #	Bed	normalized n
1	A	1.00
2	B	0.99
3	A	1.00
4	B	1.00

Relative H_2S Breakthrough Capacities

Run #	Bed	normalized n
5	A	1.00
6	B	1.02

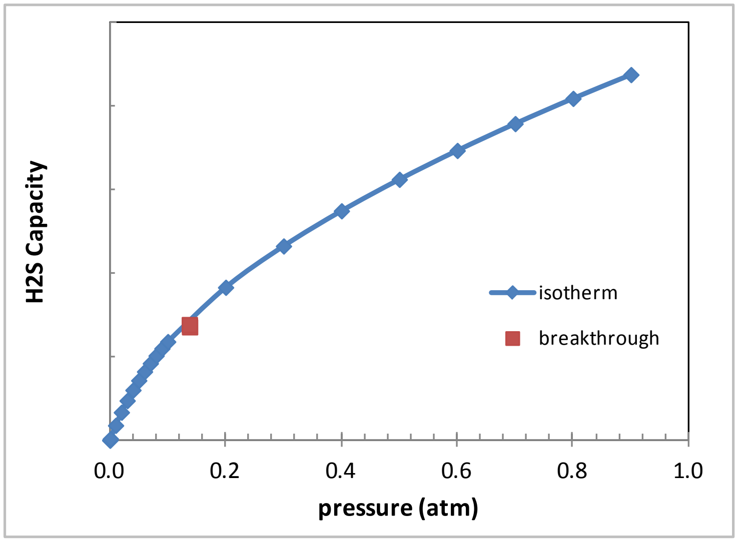


Figure 3-3. Comparison of H_2S breakthrough capacity with isotherm.

Sour Syngas Tests

The feed of sour syngas from the NCCC gasifier to our system was started on 8 October 2014 at 10 AM. The sour syngas passed through the guard beds, where organic tars were adsorbed. The beds were arranged in a lead/lag order, with sour syngas first passing through a first bed (Bed 1) and then a second bed (Bed 2). Both beds were thermally controlled to ~30^{o}C via external heat tapes, and the gas pressure was between 100 and 125 psig.

The raw syngas nominally consisted of 13% H_2, 67% N_2, 16% CO_2, 3% CO, and 1.3% CH_4, along with 300 ppm H_2S, ~10 ppm carbonyl sulfide (COS), 30-100 ppm C_2H_6, and various levels of organic tar components. The syngas was also saturated with water at the upstream knockout temperature of ~50^{o}F (10^{o}C).

The first goal was to saturate the guard beds with sour syngas so all components except for tar species passed through the beds. The light components of the syngas feed (H_2, CO, CH_4) passed through the guard beds relatively quickly. Carbon dioxide had saturated both beds after ~6 hours of gas flow. Water also showed up in the guard bed effluent gas within the first day of being onstream.

Saturating the beds with H_2S proved to be much more difficult. Air Products' previous experience with the activated carbon material at the Energy and Environmental Research Center (EERC) suggested that breakthrough of H_2S from the NCCC guard beds would take 2-5 days. In this case, the breakthrough time was much longer. This was partly due to the much lower-than-expected H_2S level in the feed gas (~300 ppm compared to 2400 ppm). Changes were made to promote quicker saturation – the lag bed was eliminated so H_2S saturation of only Bed 1 was required, and the syngas feed flow rate was increased to the maximum level allowed by the mass flow controller (40 slpm). After eight days on sour syngas, however, it was decided to accelerate H_2S breakthrough even further by feeding 0.5% H_2S/He cylinder gas to the unit at atmospheric pressure. This approach loads the adsorbent with H_2S much more rapidly than via the sour syngas while keeping the partial pressure of H_2S relatively constant (0.003 atm versus 0.005 atm). The approach was successful. After draining 2¼ gas cylinders (675 scf gas) through Guard Bed 1, small amounts of H_2S finally began to breakthrough. Sour syngas flow was resumed, and sour, tar-free syngas started to pass through the guard bed on 21 October.

Summation of all of the sour syngas and cylinder gas fed to the guard bed yielded an effective H_2S capacity of 0.57 mmole/g (~ 2 wgt %). This data point is plotted in Figure 3-4 along with the estimated H_2S capacity taken from Air Products' previous tests at EERC and a typical H_2S/carbon (coal-based) adsorption isotherm. Extrapolation of the EERC capacity with the trend of the isotherm leads one to estimate a H_2S capacity of ~0.05 mmole/g at the conditions of the NCCC guard bed (2.3 Torr H_2S). The observed value of 0.57 mmole/g is more than 10X greater than expected and led to the difficulty in saturating the bed quickly. The most likely explanation for this behavior is that there is more than just H_2S physi-sorption occurring on the carbon surface. Oxygen functionalities have been shown in the literature to provide reaction sites for H_2S decomposition to H_2O and sulfur. It is possible that this chemistry is more prevalent on the wood-based activated carbon that was used here, and this leads to 1) essentially monolayer coverage of the surface regardless of the H_2S level in the gas, and 2) formation of a rectangular isotherm shape. In the future, H_2S isotherm measurements should be conducted with the wood-based activated carbon to confirm this theory.

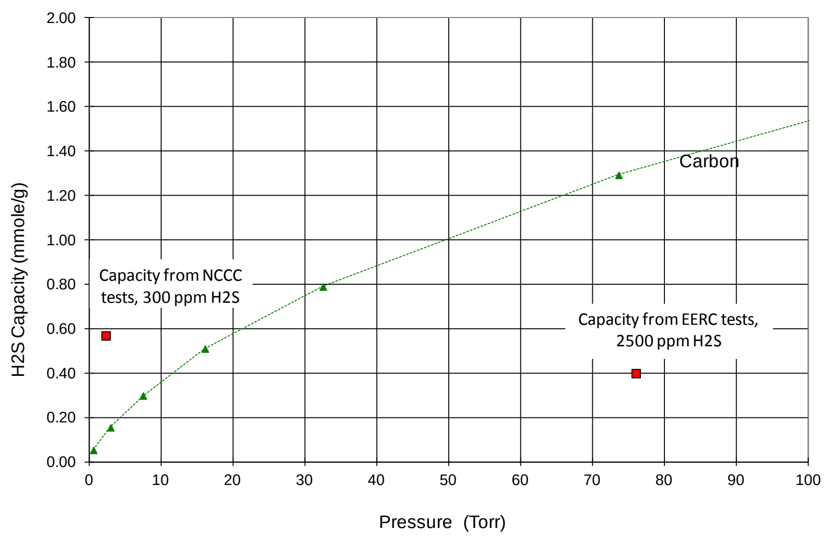


Figure 3-4. Comparison of typical H_2S/carbon isotherm with estimated capacities from guard bed breakthrough.

Throughout the H_2S saturation process and during subsequent PSA cycle operation, the progression of organic tars through Guard Bed A was monitored by measuring the benzene level in the inlet and outlet syngas. Benzene was chosen as it was by far the most prevalent species and was also one of the lower molecular weight species. It would have a relatively quick breakthrough time compared to the larger substituted aromatics and naphthalenic compounds. The inlet benzene level to Bed A was typically ~300 ppm (molar basis). No evidence of benzene, or any other tar species, was ever detected in the Guard Bed A effluent gas. The activated carbon does a great job of eliminating tar species from the sour syngas stream.

PSA Tests

The tar-free, sour syngas exiting the guard bed was compressed to ~330 psig, cooled against air, knocked out, and fed to the PSA skid. The PSA operated for a total of 12 days, or 1460 PSA cycles, on sour syngas. In addition to processing sour syngas, the PSA was operated for another 560 cycles on syngas that did not contain H_2S (before H_2S breakthrough occurred in the guard beds). The guard beds accepted sour syngas for a total of roughly 625 hours (26 days).

In all cases, the PSA operated under a simple Skarstrom cycle consisting of high-pressure feed, countercurrent depressurization to approximately atmospheric pressure, countercurrent purge with product gas at atmospheric pressure, and countercurrent repressurization with product gas. Cycle timing and typical pressures are listed in Table 3-2. The feed and gas purge pressures were set by adjusting back pressure regulators, and the syngas feed flow rate was delivered by mass flow controller at 12-19 slpm. The purge gas and repress gas flows were set to predetermined levels by appropriate setting of needle valves. The feed and purge flows were adjusted to yield a desired average level of H_2S in the product gas that was continuously measured with the galvanic H_2S analyzer. Once the flows and pressures were set, the PSA cycle was continuously executed so the process would approach a cyclic steady state, where the dynamic bed pressure/temperature and gas flows and compositions exiting the PSA are the same from cycle to cycle. Once at a cyclic steady state operating point, the performance of the PSA was characterized by measuring the inlet and outlet gas flow volumes and the average composition of the effluents. Overall and component mass balances were then determined, as well as the H_2 recovery (moles of H_2 in the product gas divided by moles of H_2 fed to the PSA) and CO_2 and H_2S rejection (moles of CO_2/H_2S in the purge and depress gas effluent streams divided by the moles of CO_2/H_2S fed to the PSA). Calculations were performed to determine feed loading, which provides the amount of feed gas processed per cycle divided by the total column volume (mlbmoles/ft^3)

Table 3-2. PSA cycle step details.

Step	Time (min)	Pressure (psig)
Syngas Feed / make Product	6	300
Countercurrent Depress 1	1	300 -> 150
Countercurrent Depress 2	1	150 -> 5
Countercurrent Product Purge	1	5
Countercurrent Repress	3	5 -> 290

Five separate PSA operating points were characterized, two with syngas without H_2S (MB1, MB2) and three with sour syngas (MB3-MB5). Typical bed temperature and pressure profiles for one of these operating points, MB4, are illustrated in Figure 3-5. The bed temperatures vary from about 15^{o}C to 40^{o}C during the cycle. Pressures and temperatures are consistent between beds A and B. Performance results derived from the operating points are listed in Table 3-3.

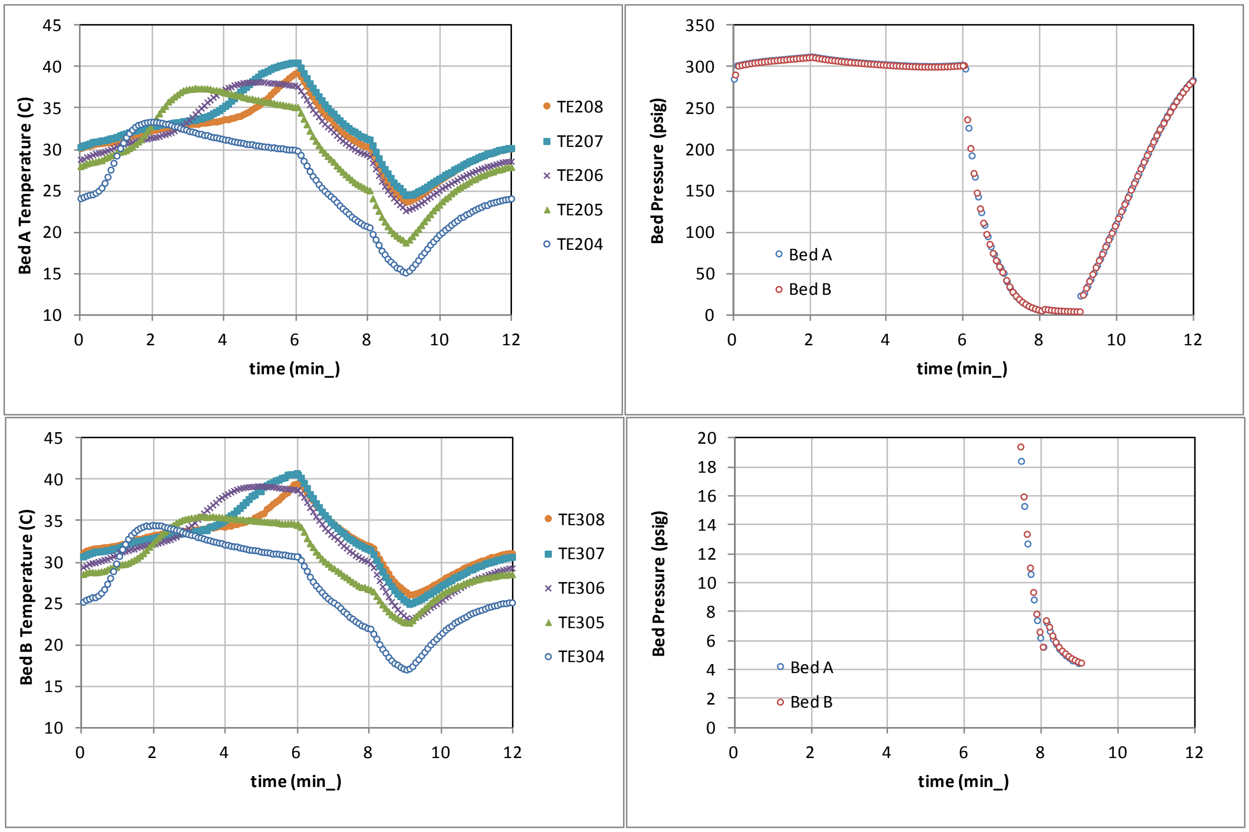


Figure 3-5. Temperature and pressure histories at cyclic steady state, MB4.

Table 3-3. Performance results from Sour PSA operating points.

MB1, In/Out		H2 recovery	58.9 %		feed	prod	dep/prg
total	1.02		70.8	H2	14.6	17.0	8.9
H2	1.14	H2S rejection na	%	N2	65.4	77.3	60.0
N2	0.96	CO2 rejection	99.6 %	CO	4.5	4.5	2.9
CO	1.23			CO2	14.6	0.1	27.3
CO2	1.12	0.11% CO2 in product		C1	0.83	1.09	0.86
C1	0.87			C2 ppm	79	0	148
C2	1.12			H2S ppm	0	0	0
H2S	na			COS ppm	0	0	0

MB2, In/Out		H2 recovery	67.1 %		feed	prod	dep/prg
total	1.00		71.0	H2	13.1	16.2	8.5
H2	1.04	H2S rejection na	%	N2	68.7	78.4	56.6
N2	1.01	CO2 rejection	98.5 %	CO	2.2	3.9	2.8
CO	0.66			CO2	15.2	0.4	31.4
CO2	1.06	0.41% CO2 in product		C1	0.68	0.99	0.81
C1	0.75			C2 ppm	60	0	114
C2	1.17			H2S ppm	0	0	0
H2S	na			COS ppm	0	0	0

Table 3-3. Performance results from Sour PSA operating points (cont.).

MB3, In/Out					feed	prod	dep/prg
total	1.01	H2 recovery	72.1 %				
			71.5	H2	13.4	16.6	9.3
H2	0.99	H2S rejection	99.7 %	N2	68.5	78.5	56.0
N2	1.00	CO2 rejection	95.37 %	CO	2.2	2.9	1.5
CO	0.98			CO2	15.2	1.2	32.6
CO2	1.08	1.4 ppm H2S in product		C1	0.72	0.85	0.51
C1	1.02			C2 ppm	38	9	86
C2	0.93			H2S ppm	256	1.5	610
H2S	1.02			COS ppm	<10	0	19

MB4, In/Out		H2 recovery	78.1 %		feed	prod	dep/prg
total	1.02		72.8	H2	12.0	15.9	8.2
H2	0.95	H2S rejection	99.3 %	N2	69.2	78.6	55.0
N2	1.02	CO2 rejection	93.94 %	CO	2.1	2.6	1.8
CO	0.92			CO2	15.7	1.6	33.8
CO2	1.09	2.5 ppm H2S in product		C1	1.03	1.25	1.07
C1	0.89			C2 ppm	49	15	104
C2	0.98			H2S ppm	212	2.5	512
H2S	1.05			COS ppm	<10	0	10

MB5, In/Out		H2 recovery	74.6 %		feed	prod	dep/prg
total	1.02		73.8	H2	13.0	16.0	9.0
H2	0.99	H2S rejection	98.5 %	N2	68.0	78.0	54.8
N2	1.00	CO2 rejection	91.90 %	CO	1.9	2.8	1.3
CO	0.87			CO2	16.3	2.2	34.2
CO2	1.15	6 ppm H2S in product		C1	0.86	1.15	0.62
C1	0.93			C2 ppm	57	19	109
C2	1.08			H2S ppm	249	6	558
H2S	1.19			COS ppm	<10	0	9

Overall mass balances were within 2%; component balances were usually within ~10%, although some outliers were also observed. The micro-GC used for the syngas component analysis appeared to be well calibrated when it was checked throughout the test period. Some of the component discrepancy could be due to ineffective mixing in the purge or product gas tank, so that an average composition was not measured.

The PSA rejected nearly all of the H_2S (>98.5%) and >90% of the CO_2 in the feed gas to the depress/purge effluent gas. Ethane and COS also preferentially partition to the depress/purge effluent gas, while N_2, CO, and CH_4 tend to partition to both the product gas and waste gas. Product gas purity for the sour syngas cases was maintained at <6 ppm H_2S even though the feed level was ~300 ppm. As expected, attainment of higher-purity product gas was achieved with reduced feed loading.

A check of the level of trace sulfur components in the feed and product gases from the PSA was made by capturing samples of these gases and submitting them for analysis with NCCC's trace sulfur analyzer. The results are listed in Table 3-4. There was no CS_2 observed in the feed or product gas. The COS levels were ~7.5 ppm in the raw syngas and 6 ppm in the feed gas to the Sour PSA unit. None of these components were evident in the PSA product gas.

Table 3-4. Trace sulfur analysis results.

Sample Name	COS	CS_2
Raw syngas, 11/2/14, A	7,42	0
	7.32	0
Raw syngas, 11/2/14, B	7.33	0
	7.66	0
	7.24	0
Guard bed effluent, A	5.67	0
	6.06	0
Guard bed effluent, B	5.72	0
	5.79	0
PSA product, B	0	0
	0	0
PSA product, A	0	0
	0	0

Hydrogen recovery was calculated in two ways: first from the measured amount of H_2 in the product, and then from the difference between the amount of H_2 in the feed gas versus the depress/purge effluent gas. If the H_2 mass balance is perfect, these two approaches will yield the same number. The results show some variation in the calculated values. Generally, the H_2 recovery is in the high 60% to mid-70% range.
One of the key goals of this test was to determine if the PSA performance was stable after the adsorbent was exposed to sour syngas. As a check, the PSA was operated at the end of the campaign with the same process setpoints that were used at the beginning of the campaign (MB3). The PSA system yielded essentially the same product composition (within experimental error), indicating that the adsorbent characteristics remained stable through the test period (Table 3-5).

Table 3-5. PSA performance at beginning and end of campaign.

Product gas composition	Condition at start of campaign	Condition at end of campaign
H_2, %	16.6	16.3
N_2, %	78.5	78.3
CO_2, %	1.2	1.2
H_2S, ppm	1.4	1.3-1.5

Final Breakthrough Tests

After the campaign, another set of breakthrough runs was conducted to see if there was any change in the CH_4 and H_2S adsorption capacities. The PSA columns were purged with N_2 overnight before the tests. The results are listed in Table 3-6 and Figure 3-6. The CH_4 capacities were ~3% lower than the earlier measurements. This change is rather low and in the range of experimental error for measuring the breakthrough of a weakly adsorbed component like CH_4. There was no change in the H_2S capacities. These results suggest there is no appreciable adsorbent decline and support the observation of stable cyclic PSA performance.

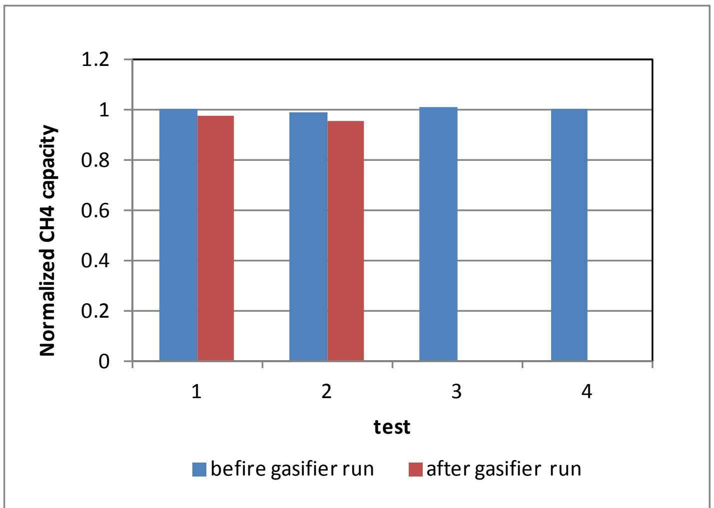
Table 3-6. Summary of final breakthrough capacities.

CH_4 Breakthrough Capacities

Run #	Bed	normalized n
7	A	0.97
8	B	0.95

H_2S Breakthrough Capacities

Run #	Bed	normalized n
9	A	1.00
10	B	1.02



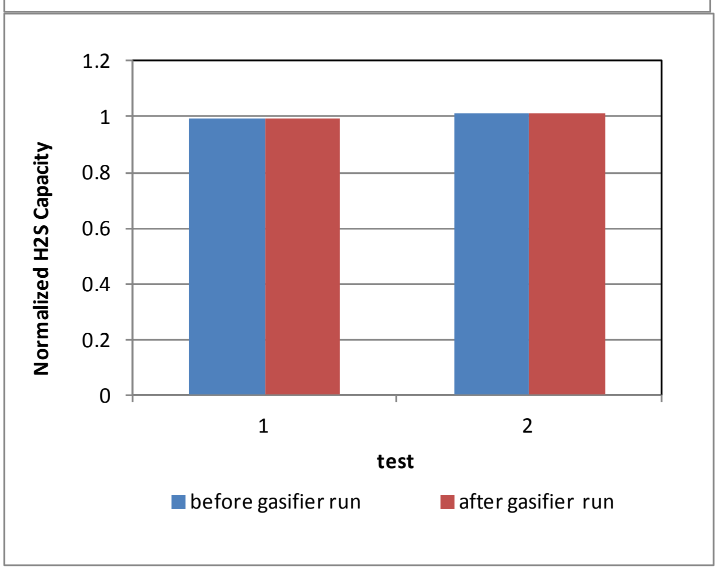


Figure 3-6. CH_4 and H_2S breakthrough capacities obtained before and after gasifier tests.

Postmortem Analysis of PSA Adsorbent Samples

When the campaign was completed, both Sour PSA columns were purged overnight with ~7 slpm of N_2 at 30°C and 2 psig. The beds were isolated, removed from the test unit, and shipped back to Air Products. Adsorbent samples were extracted from both columns within a ventilated hood in case there were significant levels of sulfur species on the adsorbent. A Tygon® tube fitted with a porous plug was inserted into the feed end of the column. A vacuum was applied to the tube, which sucked up about one inch of adsorbent from the column. The collected adsorbent was dropped into a sample vial by stopping the vacuum. In this way, samples of the column were taken from the feed end to product end of the bed.

The adsorbent samples appeared slightly tan in color after exposure, and the color was very uniform across the bed length. This is in contrast to previous tests conducted at EERC where an orange-yellow color appeared at the feed end of the bed, which quickly dissipated at the product end. There was no observable color difference between the Bed A and Bed B samples.

Adsorbent samples withdrawn from column positions roughly 4" (A4), 11" (A11), and 20" (A20) from the feed end of Bed A were submitted for postmortem testing. Similar samples from Bed B (B4, B11, and B20) were also submitted, along with a sample of fresh adsorbent.

XRF analysis was conducted to evaluate the composition of the samples. They were ground and then analyzed on an Axios WDXRF spectrometer using sample cups covered with Prolene® film in a He atmosphere. Separate qualitative scans were run to confirm the presence of all elements detected. Elements were identified using the semi-quantitative program Omnian, which estimates the concentration of all elements detected in a full scan (Na to U). The detection limit for the instrument was 0.002 wgt %. The results are listed in Table 3-6.

Table 3-6. XRF analysis of spent adsorbent samples (values are weight %).

species	fresh	B4	B11	B20	A4	A11	A20
SO3	0.16	0.16	0.18	0.24	0.22	0.23	0.21
TiO2	0.055	0.043	0.061	0.092	0.065	0.078	0.061
Fe2O3	0.04	0.04	0.044	0.065	0.055	0.073	0.051
Cl	0.029	0.03	0.026	0.033	0.032	0.031	0.026
CaO	0.015	0.024	0.022	0.02	0.016	0.017	0.019
ZrO2	0.012	0.013	0.013	0.012	0.016	0.014	0.012
CuO	0.004	0.007	0.004	<0.002	<0.002	<0.002	<0.002
ZnO	<0.002	0.002	<0.002	0.01	0.005	0.005	0.002
NiO	<0.002	<0.002	0.005	0.008	<0.002	0.005	<0.002
K2O	<0.002	<0.002	0.005	0.006	0.007	0.006	0.004
SrO	<0.002	<0.002	0.002	0.002	0.003	0.002	<0.002

The levels of detectable species on the exposed and fresh samples were similar, indicating that there is no evidence for accumulation on the exposed adsorbent.

Surface area and pore volume were measured by conducting N_2 adsorption measurements at 77 K. BET analysis yielded estimates of the adsorbent surface areas and pore volumes (Table 3-7). The surface area of the exposed samples is similar to the fresh sample, and the pore volume of the used samples appears to slightly increase. There is no indication of adsorbent degradation due to sour syngas exposure.

Table 3-7. Summary of N_2 surface area and pore volume for spent adsorbent samples.

Location	Relative BET Surface Area (m^2/g)	Relative N_2 Pore Volume (cc/g)
Fresh	1.00	1.00
A4	1.03	1.04
A11	1.01	1.04
A20	1.03	1.04
B4	0.97	1.22
B11	1.05	1.14
B20	1.01	--

An estimate of the CO_2 adsorption capacity of the fresh and used samples was evaluated by conducting TGA experiments. The TGA unit monitors the mass of the adsorbent sample as it is exposed to various gas/temperature conditions. The sample was loaded in the TGA and first heated to 200°C in N_2 for 30 minutes. It was then cooled and equilibrated at 40°C. The gas was switched to CO_2 and then N_2 to evaluate the mass change. From these tests, the weight change (and eventually adsorption capacity) under N_2 and under CO_2 could be evaluated. Results are listed in Table 3-8.

Table 3-8. Summary of TGA N_2 and CO_2 capacities of spent adsorbent samples.

capacities	Fresh	A4	A11	A20	B4	B11	B20
normalized N_2 capacity	1	1.07	1.00	1.02	1.01	0.95	1.03
normalized CO_2 capacity	1	1.01	1.03	1.03	1.02	1.03	1.03

normalized capacity = spent sample capacity / fresh sample capacity

The N_2 capacities are similar among all of the samples. The CO_2 capacities of the exposed materials are 1-3% higher than the fresh material. The results reinforce the conclusion that no adsorbent degradation has occurred.

A mass spectrometer was used in the TGA vent gas line to measure the components desorbed from the spent adsorbent samples when they were heated to 400°C. In tests of Samples A1 and A20, only water and CO_2 were observed. This indicates that the N_2 purging conducted at NCCC was sufficient to desorb the sour syngas components (CO, CH_4, C_2H_x, H_2S) from the adsorbent. Water and CO_2 were likely adsorbed onto the adsorbent during the column sampling process.

Postmortem Analysis of Guard Bed Adsorbent Samples

The guard beds were used to remove organic tar compounds from the raw sour syngas. The activated carbon used in these beds was also removed after the campaign and tested to determine if the adsorbent properties had changed.

Only one Bed 1 saw raw syngas during these tests. Before the carbon was unloaded, this bed was depressurized and countercurrently purged with hot N_2 for a period of eight hours. Vent gas was directed to the NCCC waste gas line and thermal oxidizer. Although hot N_2 was available at ~150°C, the relatively low flow resulted in the internal temperature near the bottom of the carbon bed reaching a maximum of only 50°C. Electrical heating cables on the outside wall of the vessel were used to increase the wall temperature to 80°C, but this had little effect on the internal temperature. After eight hours of N_2 purging and heating, flowing N_2 was used to cool the bed to ambient conditions.

The carbon adsorbent was removed from the beds by simply opening a ¾" NPT nozzle at the bottom of the tank and collecting the flowing adsorbent in a bucket. There was no noticeable odor from the carbon and it easily flowed through the nozzle. It appeared that the thermal regeneration was enough to remove most of the volatile species, and thankfully there were no sticky adsorbed tar components to interfere with the fluidity of the adsorbent.

The first amount of carbon to exit was from the feed end of Bed 1, so a sample was taken (and labeled "feed"). Additional samples were taken as the carbon adsorbent was drained from the bed and labeled as 2, 3, and 4. There was no easy way to determine exactly where these carbon samples originally resided in the packed bed. It is clear, though, that Sample 2 was closer to the feed end of the bed and Sample 4 was closer to the product end. Since carbon Bed 2 was never exposed to the raw or even sour syngas, it was simply purged with N_2 to remove syngas components like H_2, CO, and CH_4. A sample was also taken as the carbon was drained from the vessel (labeled as fresh).

Compositional analysis of the carbon samples obtained by XRF is presented in Table 3-9. There is definitely an accumulation of sulfur species on the samples taken from Bed 1, with the sulfur level higher at the feed end of the bed. Assuming an average SO_3 content of 4 wt% yields an equivalent H_2S loading of 1.7 wt%. Since the H_2S uptake estimated from mass balance was 2 wt%, this suggests that the majority of the sulfur is still on the carbon, likely as elemental sulfur. All of the other detectable species in the Bed 1 samples are at similar levels as fresh carbon taken from Bed 2. The sample compositions do not total 100% since the major component carbon is not detected in the analysis.

Table 3-9. XRF analysis of activated carbon in guard bed (values are weight %).

species	fresh	Feed	2	3	4
SO_3	0.04	5.4	3.9	4	3.3
P_2O_5	3	2.4	2.8	2.9	2.4
Na_2O	1.6	1.3	1.5	1.6	1.2
SiO_2	0.24	0.22	0.25	0.41	0.22
Fe_2O_3	0.033	0.031	0.031	0.037	0.022
CaO	0.026	0.021	0.024	0.021	0.015
Al_2O_3	0.023	0.023	0.015	0.27	0.018
Cl	0.005	<0.002	<0.002	0.006	0.005
MgO	0.009	<0.002	0.01	0.013	0.003
TiO_2	0.003	0.003	0.003	0.004	0.002
Br	<0.002	0.006	0.002	0.002	<0.002
K_2O	0.002	<0.002	0.002	0.002	<0.002
Cr_2O_3	0.002	0.002	0.002	0.002	0.002
NiO	<0.002	0.004	<0.002	<0.002	<0.002

Surface area and pore volumes measured for the carbons via N_2 adsorption measurements are listed in Table 3-10. The surface area and pore volume of the exposed carbon have dropped by up to 20%, particularly at the feed end of the guard bed. This is where the heavier tar species would be expected to adsorb. It is likely that some of these components were not effectively removed from the carbon. This is not that surprising considering the relatively low regeneration temperature that could be maintained in the bed.

Table 3-10. Summary of N_2 surface area and pore volume for spent carbon samples.

Location	BET Surface Area m2/g	N2 Pore Volume cc/g
Bed 2 Fresh	2204	1.291
Bed 1 Feed End	1778	1.049
BEd 1 Sample 2	2117	1.222
Bed 1 Sample 3	2158	1.255
Bed 1 Sample 4	2086	1.208

This idea is supported by CO_2 TGA data. The carbon sample at the feed end of Bed 1 and the carbon sample of Bed 2 that wasn't exposed to sour syngas were analyzed by this TGA method. Each was heated in N_2 at a predetermined temperature for 60 min, then cooled to 40°C and exposed to 1 atm CO_2 to measure the CO_2 adsorption capacity. The process was then repeated with a higher regeneration temperature. Regeneration temperatures of 150°C, 200°C, 300°C and 400°C were used. The measured CO_2 capacities as a function of regeneration temperature are plotted in Figure 3-7 .

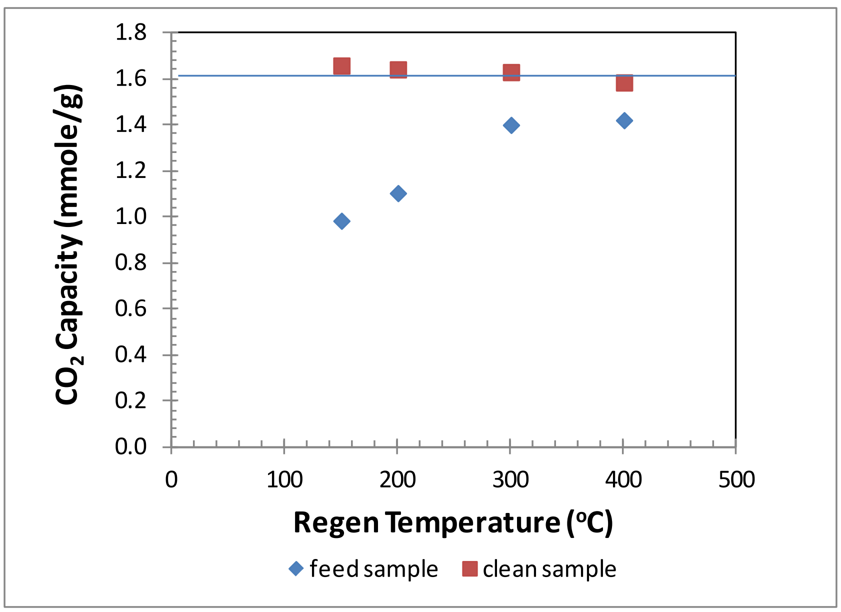


Figure 3-7. CO_2 Capacity on spent carbon samples with varying regen temperature.

The CO_2 capacity of the clean sample from Bed 2 is relatively constant at 1.6 mmole/g. The CO_2 capacity of the feed end carbon is ~40% lower when it is thermally regenerated at 150°C. Higher temperatures yield higher CO_2 adsorption capacities because more of the heavy tar species, and perhaps sulfur, are removed at those conditions. At a 400°C regeneration temperature, the CO_2 capacities are within 10% of each other. This difference could be irreversible; additional testing would be needed to verify.
For many Sour PSA applications the presence of heavy tar species is not a concern due to the aggressive conditions in the gasifier. For those applications where a guard bed is necessary, it is prudent to investigate the regenerability of the adsorbent when exposed to thermal cycles and raw sour syngas.

Process Simulations

The Sour PSA unit was built and installed at NCCC to be able to expose the PSA adsorbent to real sour syngas and evaluate the impact of that exposure. A secondary benefit was generation of some PSA performance data from the system. It is important to realize, though, that the operating conditions of the current PSA unit are very much different from an industrial unit. The current system executes a very simple cycle, contains a much shorter column and operates over a relatively long cycle time. Gas velocities in the current system are 10-15 times lower than expected in an industrial unit.

Nevertheless, efforts were taken to predict the experimental PSA performance at the various operating points with our internal adsorption process simulator (also referred to as SIMPAC in the following). The adsorption parameters used in the simulator were extracted by our standard approach. The simulations were forced to match the pressure levels in the PSA as well as the average H_2S level in the product gas. The feed gas flow rate was allowed to float in the simulations to achieve the desired level of H_2S rejection.

Parity plots for the predicted H_2 recovery and feed loading are plotted in Figure 3-8. The process simulator over-predicts the feed loading to the PSA column by about 20% and also over-predicts the H_2 recovery by up to 7 recovery points. This is not too surprising as the feed and purge gas flow rates used in these Sour PSA tests were 10-15 times lower than those used to derive the model mass transfer parameters. Mass transfer effect important at low gas velocities (e.g., axial dispersion and film resistance) are not accounted for in the simulator.

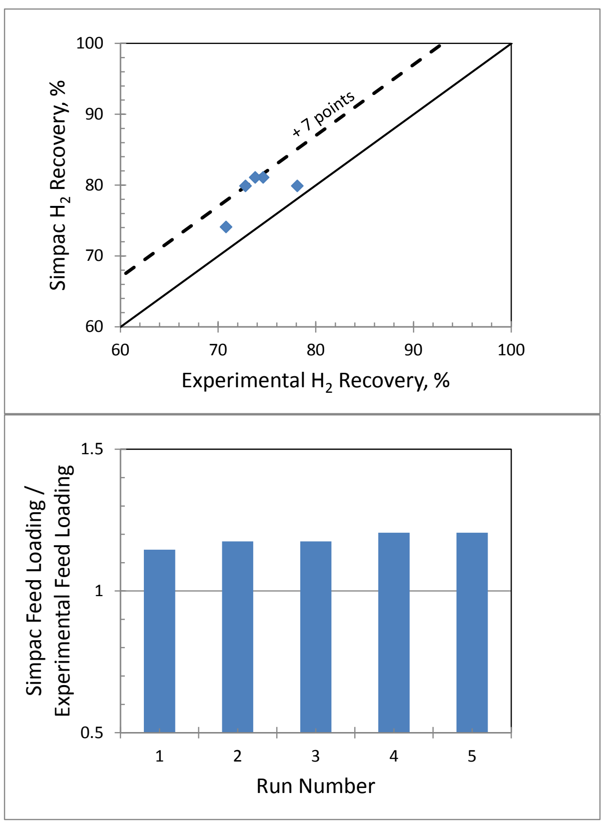


Figure 3-8. Parity plots of H_2 recovery and feed loading for model and Sour PSA experiments.

Task 4: Characterization of PSA Performance with Sulfur-Free CO/H_2 Feed Gas

In this task, Air Products' existing multi-bed PSA system (internally referred to as process development unit, or PDU) was used to experimentally validate new PSA process cycles for generating syngas suitable for methanol synthesis or hydrogen production for power generation. Since that unit could not accommodate sulfur species, it was operated with sweet syngas, but at flows and pressures relevant to industrial operation. The motivation for developing these new PSA process cycles was to reduce equipment cost by directly increasing product (H_2, H_2+CO) recovery from the PSA unit therefore alleviating the need for equipment to recover product slipping into the tail gas.

The PDU consists of a set of six packed adsorption vessels and associated equipment necessary to execute a variety of PSA process cycles. The PSA steps typically involve a high-pressure syngas feed step, various pressure equalization steps between beds, blowdown of the bed to nearly atmospheric pressure, an atmospheric pressure purge step to regenerate the beds, and finally additional pressure equalization/repressurization steps to prepare the bed for the next feed step. These steps are executed sequentially for each bed via a process logic controller. The process cycle is staggered across the beds so that the feed and effluent flows from the unit are continuous. Additional equipment is available to monitor the pressure and temperature of the packed beds, as well as characterize the composition of the product stream and flow rate of the product and waste gas streams.

Robust PSA characterization requires that the system achieve a cyclic steady state (CSS), where the dynamic performance of the beds and effluent streams is identical from one cycle to the next. At CSS, the overall performance of the PSA process can be evaluated by calculating the bed feed loading (mlbmole feed/ft^3 bed) and the hydrogen or carbon monoxide recovery at fixed product purity. The feed loading is simply the amount of feed gas to the PSA during a cycle divided by the total volume of the PSA beds. The hydrogen recovery is determined as the ratio of the amount of H_2 contained in the product gas divided by the amount of H_2 present in the PSA feed gas. A similar definition applies for CO recovery.

In this work, the PSA beds were packed with adsorbent to a final length of 25 ft. Feed gas consisting of a mixture of hydrogen, carbon monoxide, carbon dioxide, argon, methane and nitrogen was generated from gas cylinders/hydrotubes and delivered to the PSA beds by mass flow controllers. The flow rate of effluent gas (product and blowdown/purge gas) was measured with dry test meters. Composition of the product gas was determined by a combination of infrared analyzers (CO_2, CO, CH_4) and gas chromatography units (N_2, Ar, H_2). The PSA operated at room temperature (controlled to ~75^{o}F), a feed pressure of 430 psig and purge pressure of 15 psig.

Syngas purification for methanol production and power production were considered during two separate campaigns. Details of each study are described below.

Production of Syngas for Methanol Synthesis

The PSA columns were each packed with a single layer of Air Products' preferred sour gas adsorbent. An average of 1,364 g of adsorbent was packed in each column for a bulk density of 48.5 lb/ft^3.

Nine separate runs were conducted at various process flow rates or purity setpoints. Feed gas compositions are listed in Table 4-1. All runs considered a 2:1 H_2:CO feed ratio except the last one, which was adjusted to 3:1 H_2:CO.

Table 4-1. Feed gas for methanol cases.

	feed gas for PDU runs 385 - 1-8	feed gas for PDU run 385 - 9
H2/CO	2.09	3.00
H2	43.34	48.08
CO2	35.54	35.54
CH4	0.0120	0.012
Ar	0.0180	0.018
CO	20.77	16.03
N2	0.32	0.32

Tests of a PSA cycle designed to maximize the recovery of H_2 and CO were conducted. The cycle steps are outlined and described in Table 4-2. The cycle was unique because after the feed step it included a rinse step using CO_2 provided from gas cylinders. The overall cycle required 1,440 seconds to complete, or roughly 24 minutes. The PSA system was continuously cycled until the average effluent gas composition stabilized and became constant (at CSS). The feed gas flow rate was simultaneously adjusted during the approach to CSS to yield an average of 1% CO_2 in the product gas. This typically took 40-70 cycles, or roughly 15-30 hrs.

Typical bed pressures during one of the runs is illustrated in Figure 4-1. The pressure for each bed is shown as a separate line, and the similar traces indicate that the pressure histories are very consistent.

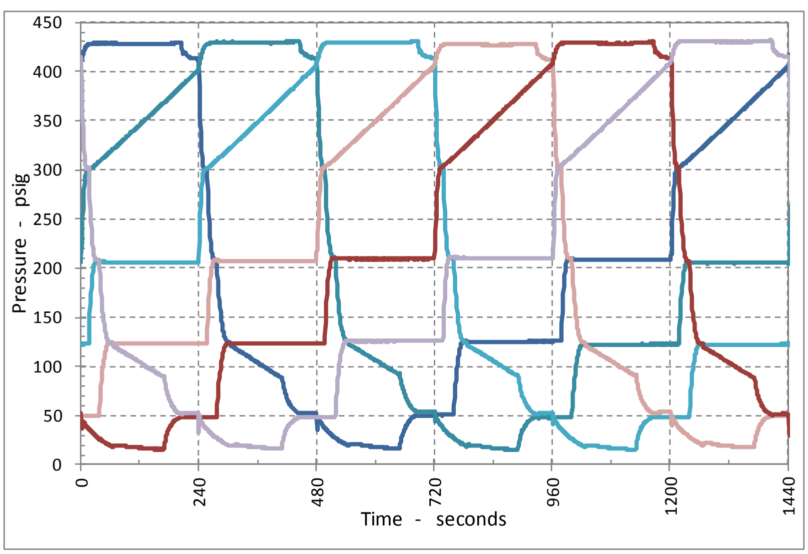


Figure 4-1. Pressure trace for PDU run 385-01-04.

A summary of the performance results from the nine experiments is listed in Table 4-3. The rinse cycle yields H_2 recoveries in the high 90% range, while CO recovery generally falls in the 80-90% range. The difference is due to the stronger adsorption strength of CO compared to H_2. The H_2/CO ratio of the product gas is always slightly more H-rich than the feed gas, also an effect of the increased CO adsorption strength. The PSA is very capable of reducing the CO_2 levels to those needed for methanol synthesis (<1%) while maintaining an appropriate CO/H_2 ratio. Inert levels (N_2, Ar) are < 0.35% and CH_4 is 100 ppm or less.

Table 4-2. Cycle table for the methanol syngas application.

col 1		feed				RNS	eq1	eq2	eq3	pr purge	eq4		blowdown			purge	eq4				eq3					eq2					eq1		repress			
col 2	eq1		repress					feed				RNS	eq1	eq2	eq3	pr purge	eq4			blowdown		purge	eq4				eq3					eq2				
col 3		eq2					eq1		repress					feed				RNS	eq1	eq2	eq3	pr purge	eq4			blowdown		pr purge	eq4				eq3			
col 4			eq3					eq2					eq1		repress					feed				RNS	eq1	eq2	eq3	pr purge	eq4			blowdown		purge	eq4	
col 5		blowdown		purge	eq4				eq3					eq2					eq1		repress					feed				RNS	eq1	eq2	eq3	pr purge	eq4	
col 6	eq1	eq2	eq3	pr purge	eq4			blowdown		purge	eq4				eq3					eq2					eq1		repress					feed				RNS

feed = high pressure syngas feed step / make product
RNS = high pressure cocurrent CO_2 rinse
eqx = bed to bed pressure equalization step x
pr purge= provide purge
blowdown = countercurrent bed depressurization to ~ atmospheric pressure
purge = receive purge at ~ atmospheric pressure
repress = countercurrent pressurization with product gas

Table 4-3. Results from PDU tests; methanol cases.

experiment -->	385-01-01	385-01-02	385-01-03	385-01-04	385-01-05	385-01-06	385-01-07	385-01-08	385-03-01
feed (scfh)	81.07	81.96	92.66	89.79	74.04	82.03	87.71	94.74	83.66
CO2 rinse (scfh)	47.50	47.50	0.00	47.50	47.50	23.90	23.90	47.50	47.50
H2 recovery (%)	93.8	95.1	96.8	97.9	97.9	98.5	101.2	95.9	98.2
CO recovery (%)	86.8	85.8	83.7	91.3	73.9	75.9	77	88.1	81.6
H2/CO in product	2.26	2.33	2.42	2.25	2.78	2.73	2.76	2.31	3.62
feed temperature (degF)=	77	77	77	77	77	77	77	77	77
material balance error (%)	0.89%	1.38%	1.20%	1.93%	2.00%	2.18%	2.10%	0.72%	2.02%
normalized feed loading	100.0	101.1	114.3	110.8	91.3	101.2	108.2	116.9	103.2
product composition									
H2 in product (%)	68.37	68.99	69.85	67.32	73.16	72.70	72.53	68.75	77.39
N2 in product (%)	0.278	0.300	0.317	0.298	0.280	0.301	0.306	0.281	0.265
CO2 in product (%)	1.04	1.07	0.98	2.48	0.21	0.32	0.85	1.21	0.95
CO in product (%)	30.29	29.61	28.83	29.89	26.34	26.66	26.30	29.73	21.38
CH4 in product (ppm)	86	94	96	100	52	93	98	108	91
Ar in product (ppm)	115	125	140	115	100	122	118	125	114

As in any PSA, higher product purity (i.e., less CO_2) can be achieved by decreasing the feed rate to the unit (lower feed loading). This is demonstrated in the data (Runs 385-01-02, -01-04, and -01-05). Run 385-01-08 gives the best performance in the group with 96% H_2 recovery, 88% CO recovery, and the highest feed loading measured in these tests.

The process simulator was used to predict the performance of these PDU experiments. The pressure history of the beds was matched relatively well, as illustrated in Figure 4-2. Specific results for experiment 385-01-08 in Table 4-4 show that the simulator over-predicts the H_2 recovery and the feed loading. The product composition is reasonably close, although the level of inerts (N_2 and Ar) is over-estimated by SIMPAC.

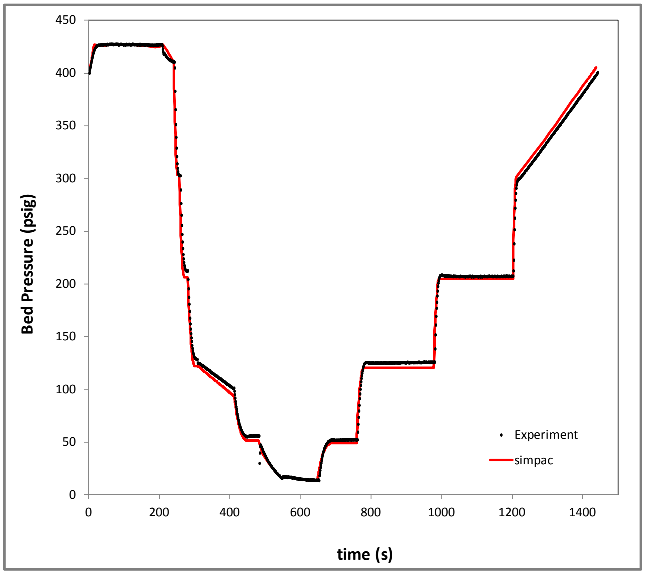


Figure 4-2. Bed pressures for cyclic PSA bed and simulation (methanol case).

Table 4-4. Comparison of PDU experiment 385-01-08 with SIMPAC prediction.

experiment -->	385-01-08	SIMPAC
feed (scfh)	94.74	103.28
CO2 rinse (scfh)	47.50	47.50
H2 recovery (%)	95.9	97.5
CO recovery (%)	88.1	88.52
H2/CO in product	2.31	2.30
feed temperature (degF)=	77	77
normalized feed loading	100.0	109.0
product composition		
H2 in product (%)	68.75	68.52
N2 in product (%)	0.281	0.486
CO2 in product (%)	1.21	1.20
CO in product (%)	29.73	29.75
CH4 in product (ppm)	108	129
Ar in product (ppm)	125	260

Overall results for the experimental and model recoveries and feed loadings can be found in the parity plots of Figure 4-3. Recoveries are generally predicted within 3 points of the measured values, and the SIMPAC feed loadings are consistently about 10% high. Since the recovery predictions were well-distributed across the error range in the parity plot, there was little potential to improve the fit to the data by manipulating the adsorption parameters.

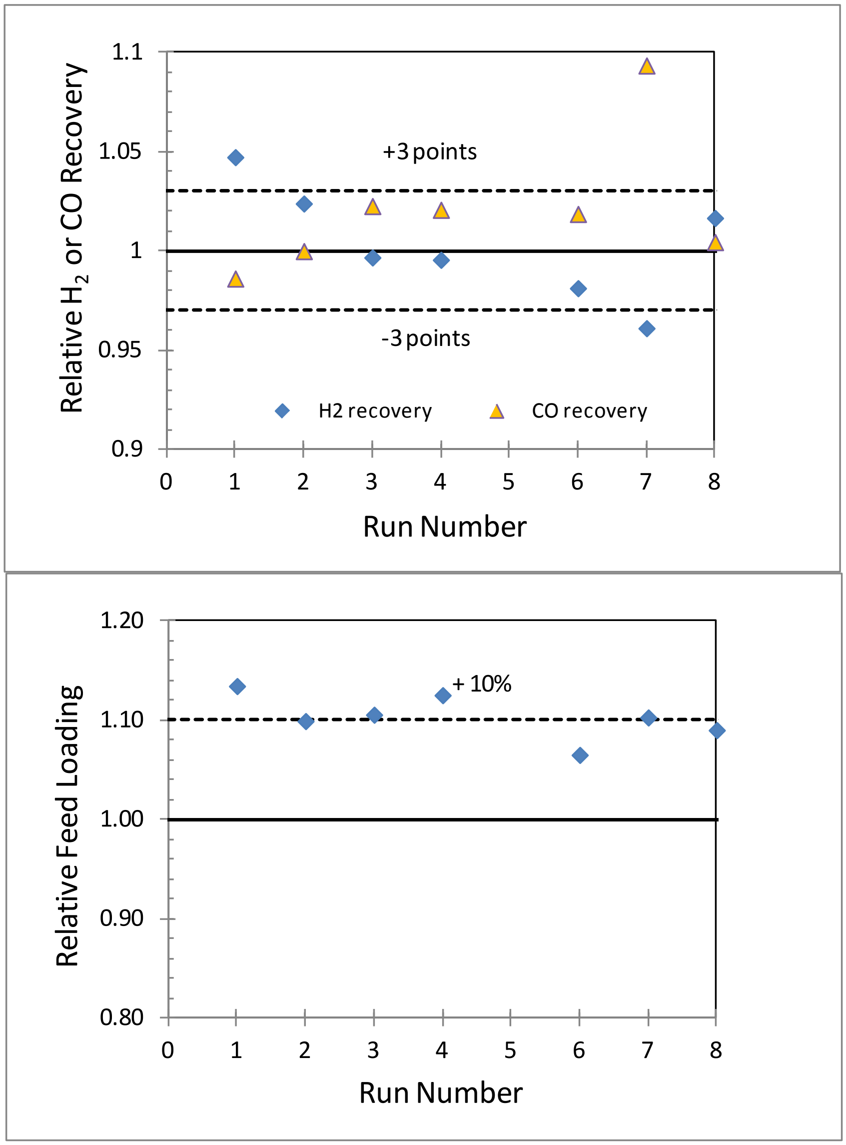


Figure 4-3. Parity plot for syngas PDU data and model predictions.

Production of Hydrogen for Power Generation

For this test, the six columns were packed with 10 ft of Air Products' preferred sour gas adsorbent along with 8 ft of activated carbon and 7 ft of zeolite. The additional adsorbent layers were added to more effectively remove CO_2, CH_4 and CO from the syngas. Six separate runs were conducted. In an industrial application the gasifier syngas would pass through a water/gas shift unit to maximize the H_2 content of the feed gas, so the feed gas blend in the PDU tests was modified to the composition listed in Table 4-5.

Table 4-5. Feed gas for ammonia PDU runs.

	feed gas for PDU Ammonia runs
H2	53.33
CO2	46.00
CH4	0.0107
Ar	0.0168
CO	0.39
N2	0.25

The PDU was operated in a manner similar to the syngas runs, except now the CO content of the product gas was the control variable. It was typically set to 1 ppm, although tests with higher and lower levels were also conducted. The process cycle was also changed to one that could produce a diluted H_2/N_2 product blend needed for power production. Hydrogen dilution is necessary for controlling internal gas turbine temperatures to reasonable levels. This same diluted syngas could also be a feedstock for ammonia production. The various steps are illustrated in Table 4-6. It included two modified steps where a high-pressure nitrogen rinse was combined with pressure equalization steps, along with a low- pressure purge step conducted with nitrogen rather than product gas. Both changes tend to reduce the loss of H_2, increase its recovery in the product gas, and add N_2 to the H_2 product stream.

A summary of the results is presented in Table 4-7. The recovery of H_2 was experimentally demonstrated to be very high ($\geq$97%) with this improved PSA cycle. The recovery was also shown to be relatively insensitive to the operating parameters investigated in these runs (different purge flows, rinse flows, and product purity levels). The CO content of the product gas was varied from 0.2 ppm to 20 ppm by increasing the feed rate by 7%. CH_4 was also nearly completely removed from the product gas (< 1 ppm). The major components of the product stream consisted of ~85% H_2 and 15% N_2. Further dilution of this product stream with N_2 and/or steam would result in a suitable fuel for a gas turbine in power generation applications. Similarly, a suitable ammonia feed could be easily obtained by blending additional N_2 into this stream to achieve the required 3:1 H_2/N_2 ratio.

These data were also compared with process simulations. The pressure histories from the simulations were set to track those in the experiments, and the results showed reasonable agreement (Figure 4-4). The H_2/N_2 product split is listed in Table 4-8. The simulations predict a slightly more N_2-rich product than observed in the experiments.

Table 4-6. Cycle table for power production application.

col 1	feed						eq1/R	eq2/R	eq3			eq4	blowdown				N2 prg	eq4			eq3					eq2					eq1	repress				
col 2	eq1	repress					feed						eq1/R	eq2/R	eq3			eq4	blowdown				N2 prg	eq4			eq3					eq2				
col 3		eq2					eq1	repress					feed						eq1/R	eq2/R	eq3			eq4	blowdown				N2 prg	eq4			eq3			
col 4			eq3					eq2					eq1	repress					feed						eq1/R	eq2/R	eq3			eq4	blowdown				N2 prg	eq4
col 5	blowdown				N2 prg	eq4			eq3					eq2					eq1	repress					feed						eq1/R	eq2/R	eq3			eq4
col 6	eq1/R	eq2/R	eq3			eq4	blowdown				N2 prg	eq4			eq3					eq2					eq1	repress					feed					

feed = high pressure syngas feed step / make product
eqx/R = bed to bed pressure equalization step x with simultaneous cocurrent N2 rinse
blowdown = countercurrent bed depressurization to ~ atmospheric pressure
N_2 prg = receive N_2 purge at ~ atmospheric pressure
repress = countercurrent pressurization with product gas

Table 4-7. Results from PDU tests; power production case.

name	description	feed time	feed	rinse	purge	H2 Recovery	mat bal error	Product					
								H2	N2	CO	CO2	CH4	Ar
		s	scfh	scfh	scfh			%	%	ppm	ppm	ppm	ppm
386-01-01	base case, CO = 1 ppm	210.00	84.72	72	22.3	97.2%	-0.47%	86.1	13.9	1.0	0.0	0.200	123
386-01-02	lower purge	210.00	78.67	72	18	97.7%	1.12%	87.0	13.0	0.9	0.0	0.352	103
386-01-03	higher rinse	220.00	71.3	90	22.3	97.4%	1.01%	83.4	16.6	0.9	0.0	0.310	130
386-01-04	repeat 1, modify PE4 time	210.00	83	72	22.3	97.6%	0.45%	85.6	14.4	2.1	0.0	0.220	130
386-01-05	lower CO spec. 0.3 ppm	210.00	79.85	72	22.3	97.8%	0.77%	86.6	13.4	0.2	0.0	0.250	105
386-01-06	high CO spec, 20 ppm	210.00	85.49	72	22.3	98.0%	0.51%	84.5	15.5	20.8	0.0	0.599	156
386-01-07	higher CO spec, 5 ppm	210.00	82.58	72	22.3	98.3%	0.95%	85.1	14.9	4.9	0.0	0.460	139

Table 4-8. Product composition for power production case.

	Product composition		
	y H2	y N2	y CO
	%	%	ppm
Experiment	86.1	13.9	1.0
Simulation	84.8	15.2	1.0

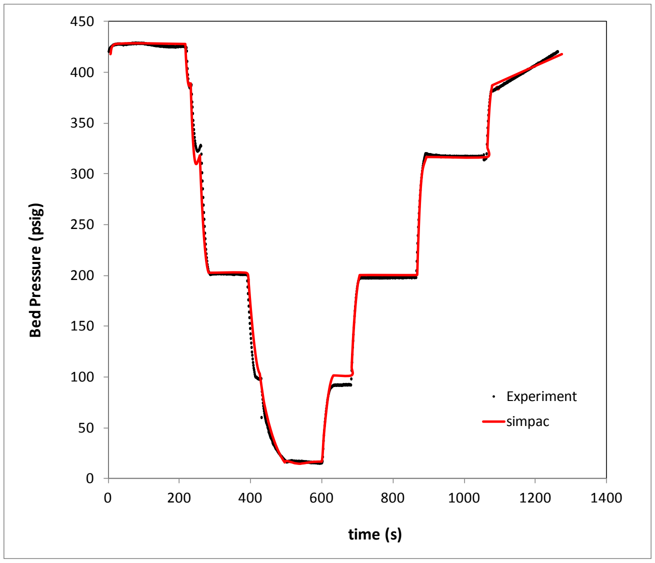


Figure 4-4. Bed pressures for cyclic PSA bed and simulation (power case).

The parity plots in Figure 4-5 show that the H_2 recovery and feed loading for the power production case are both under-predicted by SIMPAC with the current adsorption parameters. The predicted feed loading is about 10-20% lower than experimental, and the H_2 recovery is lower by roughly 4 recovery points. This discrepancy can be eliminated by adjusting the adsorption parameters for the zeolite adsorbent, but there was no need to do this in the current work because N_2 rinse cycles were not utilized in the techno-economic analysis.

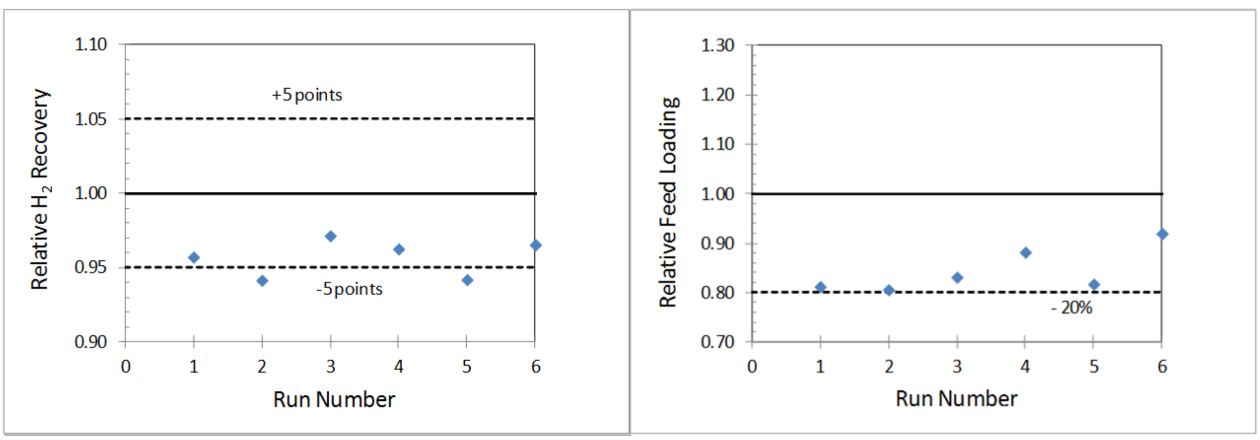


Figure 4-5. Parity plots for power production case, PDU data and model predictions.

Task 5: Techno-Economic Analyses of Sour PSA for Syngas Production for IGCC and Methanol Production

Techno-economic assessments (TEAs) of the performance of the Sour PSA technology were developed for both IGCC and methanol production following guidelines published by NETL/DOE. In both cases, a baseline or reference case using conventional technology for acid gas removal was first established. Then the TEA for a plant utilizing the Sour PSA technology was established using the same assumptions, allowing for direct comparison and estimate of the financial benefits of the new technology. Cost comparisons are made in 2011 US dollars.

The IGCC cases were built using the same plant design as in the NETL 2011 "Cost performance baseline for fossil energy plants" and the methodology described in "Cost estimation methodology for NETL assessment of power plant performance" publications. [2,4] Costs were updated to 2011 US$ using a combination of data from the NETL "Updated cost (June 2011 basis) for selected bituminous baseline cases" [3] and the Chemical Engineering Plant Cost Index evolution between 2007 and 2011 (11.5% increase).

The methanol production case was built using the same plant design and assumptions described in the NETL 2014 "Baseline analysis of crude methanol production from coal and natural gas." [8] The project finance structure was made using assumptions described in the NETL 2011 "Recommended project finance structures for the economic analysis of fossil based energy projects." [5]

In all cases, high-level heat and mass balance around the relevant units of operation were simulated in steady-state regime using Aspen Plus.® The simulation results were then used to estimate equipment size and process scale-up factor. Capital cost estimates of scaled equipment were performed using the methodology presented in the NETL 2013 "Capital cost scaling methodology" publication or using an exponential factor of 0.6 for a few equipment pieces not covered in the NETL document. [6] Internal engineering estimates were used for the capital and installation costs of new equipment related to implementation of Air Products' Sour PSA technology. Finally, fuel prices were taken from the NETL 2012 "Fuel prices for selected feedstock in NETL studies" publication. [7]

Techno-economic evaluation of Sour PSA technology for IGCC applications.

A TEA was completed for the impact on Cost of Electricity (COE) from incorporating of Air Products' Sour PSA technology into a base IGCC power plant with CO_2 capture design utilizing low-rank coal.

The 2011 "Cost performance baseline for fossil energy plants" report by NETL [2] was used to guide the methodology of estimating the overall economics of the project. This methodology was also applied to the information provided for the baseline reference plant case to insure the accuracy of this approach. Upon review of the available cases, Air Products selected the Siemens gasifier with CO_2 capture and sub-bituminous PRB coal case (referred as case S3B in the NETL report) as the baseline case for this assessment. The reference system, shown in Figure 5-1, includes a coal handling system that feeds a train of a Siemens gasifier and an air separation unit (ASU) that provides O_2 to the gasifier as well as N_2 for diluting the gas turbines fuel. The raw syngas produced by the gasifier is scrubbed and then shifted to H_2 and CO_2 before it undergoes Hg removal and is then sent to the acid gas removal (AGR) unit. The AGR unit consists of a 2-stage Selexol unit and a Claus plant with tail gas cleanup unit. The CO_2 product is compressed to pipeline specifications, and H_2 is diluted with N_2 to fuel for two F-type gas turbines in combined-cycle arrangement with a single steam turbine through a HRSG.

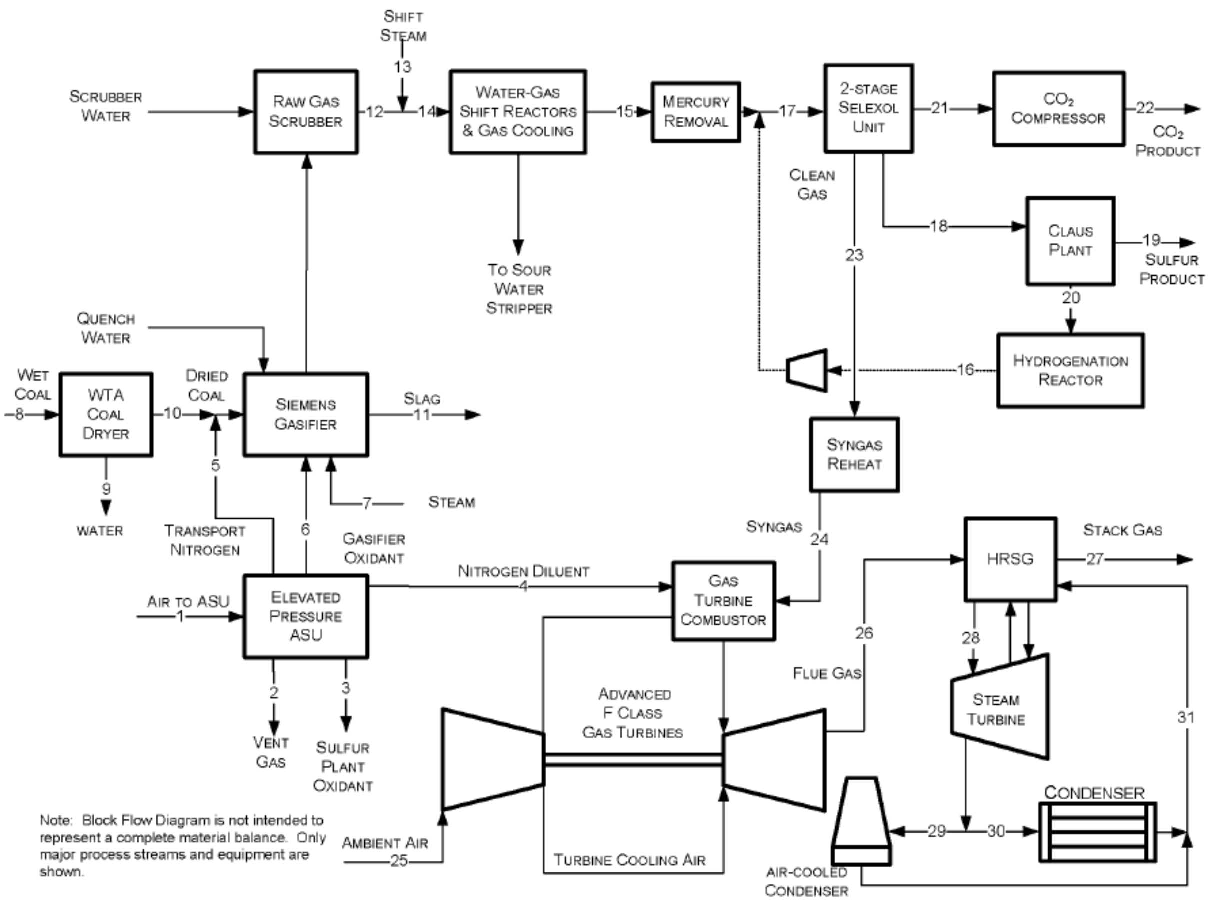


Figure 5-1. Reference IGCC process (from [2]).

For this evaluation, the gasifier size was fixed but the AGR unit was replaced with Air Products' Sour PSA technology. This resulted in the need to resize the power island and the ASU to account for the performance deviation from the reference case, but allows for the gasification island, shift and other ancillary equipment to remain the same. Both the AGR and Sour PSA systems are fully integrated into the overall plant process, which requires one to account not only for equipment replacement but also equipment size and/or performance modifications. This reduces the potential for associated with having to scale multiple unit operations costs. However, it does introduce issues with the gas and steam turbines, as that equipment is not typically engineered to order. However, the approach is useful to compare technologies relative cost on a unit operation basis with limited uncertainties.

The Sour PSA acid gas removal system and the altered steam cycle of the power island are simulated in steady state using Aspen Plus.® The gas turbines are treated as "rubber" gas turbines having the same heat rate performance (for a given volumetric heat value of the fuel) as the reference case. This is a reasonable assumption, as the gas turbine is operated near the reference-case fuel conditions.

A comprehensive economic model capable of predicting the COE has been developed with the same assumptions and economic scenarios used in the NETL report. This model includes Total Plant Cost and Initial and Annual O&M Costs, allowing for a simple estimation of the COE. Note that the economic model is capable of calculating COE from a cash flow analysis similar to that used in the DOE's Power System

Financial Model (PSFM). [9] Estimating COE from the cash flow analysis gives results consistent with those published by NETL for a non-CO_2 capture cases. However, the COE published by NETL for the CO_2 capture cases does not entirely result from a cash flow analysis, since the cost of CO_2 transport, sequestration, and monitoring (TS&M) is treated separately as an addition to the COE without TS&M cost. To ensure consistency, in this study the COE was calculated using the same method that NETL followed: using a capital charge factor for high-risk projects and adding the CO_2 TS&M cost separately. In addition, a COE number was calculated for the Sour PSA system based on the cash flow statement.

Process selection

Both the reference case and the plant using the Sour PSA technology are assumed to be located in Montana and use PRB coal as feedstock. The cost of fuels and cost of CO_2 Transport and Storage (T&S) are calculated accordingly based on NETL guidelines. [7, 3]

Reference case

The reference case used here is the so-called S3B case from the Cost and Performance Baseline for Fossil Energy Plants - Volume 3A: Low Rank Coal to Electricity: IGCC Cases [2] and consists of the following:

- Two ASU trains
- Two trains of coal drying (WTA) and dry feed system
- Three trains of gasification, including gasifiers (Siemens), a syngas cooler, and particulate removal
- Two train of syngas cleanup process
- Two trains of two stage Selexol acid gas removal system
- One train of Claus based sulfur recovery
- Two gas turbine/HRSG tandems
- One steam turbine

New case using Sour PSA Technology

The results from the experimental tests conducted in Task 3 provided a basis on which to model, design, and cost a process configuration for the TEA. The best solution for a particular gasification site depends on many variables, including type of feedstock, gasification technology employed, desired primary products (H_2, power, or syngas), impurities to be removed to satisfy downstream processes, and ultimate disposition of CO_2 and other impurities. For a coal feedstock in particular, the challenges and solutions are different, depending on high- or low-sulfur feedstock, the properties of the ash, and various levels of impurities like arsenic, lead, vanadium, mercury, chlorine, and fluorine. Lower-rank coals can be especially difficult to gasify, as they produce substantial amounts of by-products like tars that foul heat transfer surfaces and plug packed beds. Some low-rank coals contain high levels of alkali metals that, in addition to the above problems, can aggressively corrode materials of construction. The nature of the tars and alkali is specific to the particular coal and the environment under which it is gasified.

The primary objective of Sour PSA technology is to lower the cost of acid gas removal from a gasification plant. It is important to understand, however, that Sour PSA is only one component, albeit the primary enabler, of a proprietary complete downstream process that more fully accomplishes these objectives. The complete Sour PSA system is shown in Figure 5-2. The system consists of three components: 1) a PSA unit for purification of the sour syngas stream; 2) sulfur treatment, and 3) CO_2 polishing and compression. Unlike a traditional AGR system, a single waste stream (tail gas) is produced from the Sour PSA unit containing impurities along with a small amount of H_2. Process options for treatment of the tail gas will vary depending on the desired disposition of the sulfur (elemental or sulfuric acid-based) and the products of the plant (H_2, power, syngas, or syngas-based products).

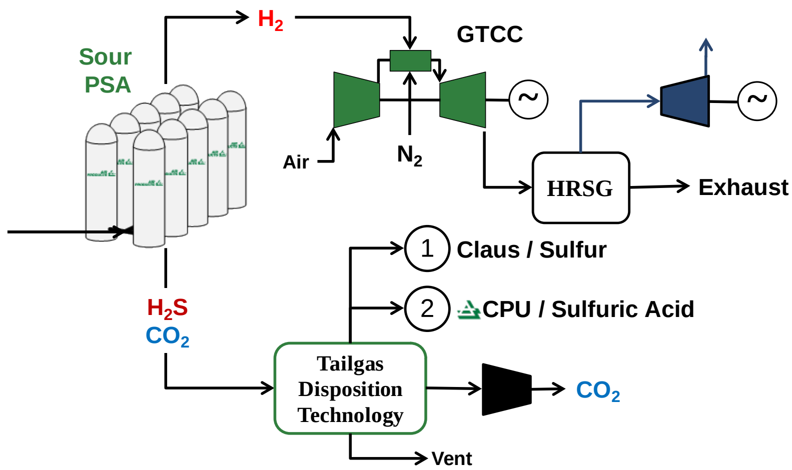


Figure 5-2. Tail gas treatment options for power applications of Sour PSA technology.

One option for a Sour PSA process configuration is the Sour PSA unit followed by a sour oxy-combustion unit and finally the CO_2 compression/purification unit (CPU), as shown in Figure 5-3. The Sour PSA is fed sour syngas and produces a high-pressure, H_2-enriched product and a low-pressure, CO_2/H_2S-rich tail gas. The oxy-combustion process is used to effectively combust flammable species in the tail gas (H_2, CO, CH_4) to CO_2 and H_2O, and H_2S to SOx and H_2O. This creates an effluent stream that contains highly enriched CO_2 with minor impurities. The heat generated from the combustion system can be used for preheating streams to a turbine in a power system, steam generation, additional reforming in a hydrogen system, or any other ancillary use of high-quality heat. The oxyfuel combustion may take place either in a once-through manner, or with cooled flue gas recycle to moderate the combustion temperature.

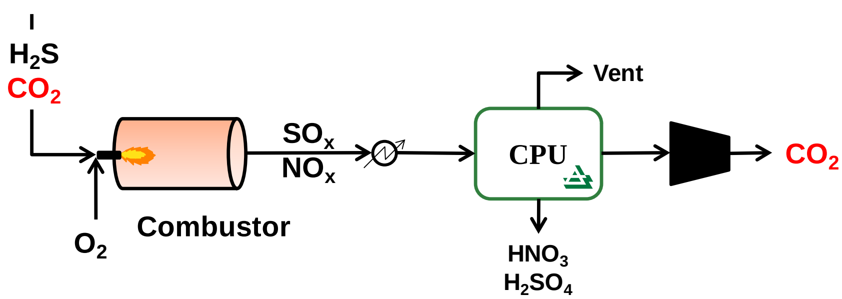


Figure 5-3. The combustion/CPU tail gas treatment option.

The sour oxyfuel combustion is accomplished by combusting the waste gas with an excess of pure O_2, in which case the combustion products will be H_2O, CO_2, SO_2, SO_3 and excess O_2. The SO_2 and excess O_2 may be removed from the CO_2 by reactive processes applied during the compression sequence. Specifically, this requires careful design of the compression system coupled with acid production reactors of appropriate size. SO_2 is removed as H_2SO_4, and NO and NO_2 are removed as HNO_3 in that system. The SO_x-free, NO_x-lean CO_2 gas may then be compressed to pipeline pressures and either stored in geological formations or used in enhanced oil recovery (EOR). Associated byproducts from the purification system are H_2SO_4 and HNO_3, which may either be saleable in the given market or disposed of in an appropriate manner. The specialized CO_2 compression/purification system, which includes SO_2, NO_x, and inert removal systems, was originally developed by Air Products for oxyfuel CO_2 capture for pulverized coal combustion power boilers. This technology is currently in the pilot stage of development at Vattenfall's oxyfuel combustion pilot plant at Schwarze Pumpe, Germany.

A second option to treat the tail gas from the PSA is to remove the H_2S prior to purification of the CO_2 product stream. Figure 5-4 shows a schematic in which H_2S and a small portion of CO_2 are removed in an acid gas enrichment (AGE) step. The AGE is generally configured in an absorber/stripper arrangement with a solvent that is selective to H_2S. MDEA is a common amine for this service. Some licensors add a promoter to the MDEA to improve the selectivity to H_2S. Others, such as ExxonMobil's Flexsorb,® have specialized sterically hindered amines. The acid gas produced from the stripper of the AGE is sent to a Claus plant for sulfur recovery. The sweetened gas is then compressed (~30 bar) and sent to an auto-thermal refrigeration/partial condensation unit, where CO_2 impurities are rejected as lights and a product CO_2 stream is formed. The resulting product CO_2 stream is then further compressed to required pipeline pressures. The light elements stream ("lights") is at pressure and rich in hydrogen and other combustibles originally rejected in the tail gas stream with low levels of H_2S and CO_2. This stream is of a sufficient pressure and heating value to send directly to gas turbine as supplement fuel with proper adjustment of the heating value with N_2 from the ASU. This creates a pathway to recover the energy stored in the hydrogen originally removed in the PSA while maintaining a high rate of CO_2 capture. This option also employs commercially available technologies to produce both sulfur and CO_2 product streams, thus reducing the overall risk associated with implementing the technology.

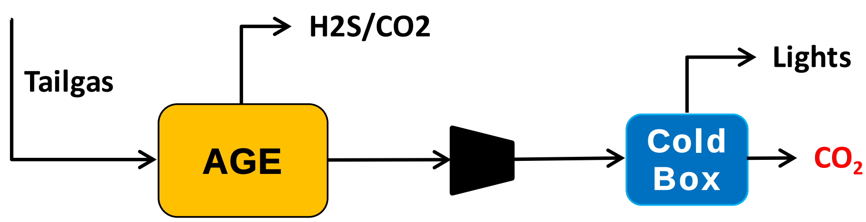


Figure 5-4. Sour PSA acid gas enrichment tail gas treatment option for elemental sulfur product disposition followed by a cold box to purify the product CO_2

Air Products has demonstrated in a previous TEA of Sour PSA technology for IGCC applications that in the case of gasification of low-sulfur-content coal like PRB, disposing of the sulfur from the PSA tail gas using an acid gas enrichment step was the more favorable option. [1] Therefore, this is the only option considered in this TEA.

Process simulation

In order to allow a direct comparison between the plant using the Sour PSA technology and the reference case S3B [2], both plant process where simulated. More specifically, the portion of the process that would be impacted by implementation of the Sour PSA technology was simulated, i.e. downstream of Hg removal and the power island.

Reference process cycle

To compare the performance of Air Products' Sour PSA technology to the reference case using Selexol as described in the NETL report (S3B case)[2], a complete understanding of the power island is necessary. Indeed, the power island's performance is directly impacted by the choice of technology used for the acid gas removal. The NETL report provides only limited information on the steam cycle. Since it is critical the different cases are compared on the same basis, Air Products simulated the process of the power island for the reference case with some assumptions. Figure 5-5 shows a simplified PFD of the simulated process that includes these assumptions. The corresponding heat and mass balance summary is provided in Appendix Table A5-1.

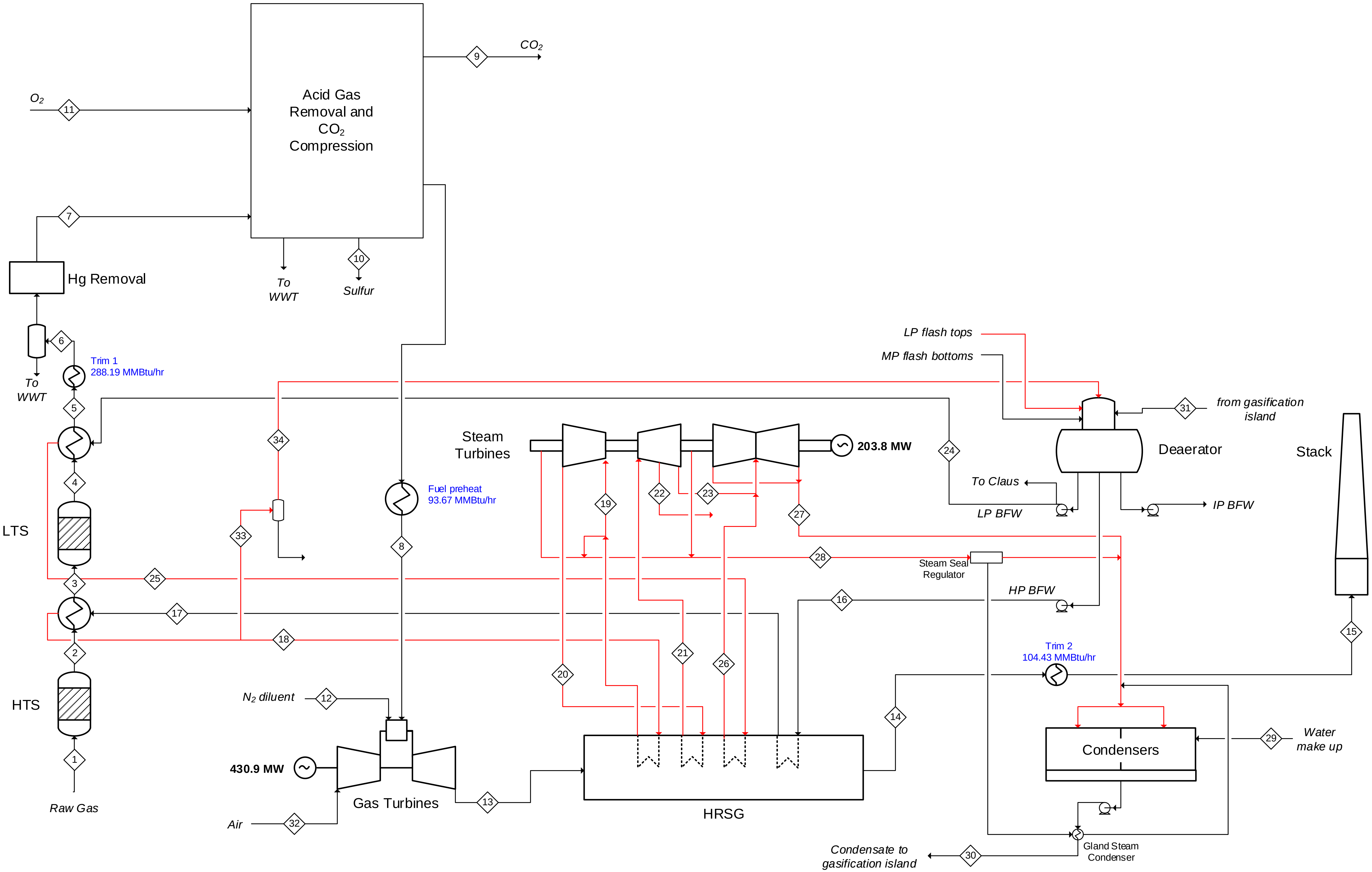


Figure 5-5: Simplified PFD of the simulated reference case process.

Comparing this simulated process with the information from the NETL report suggests that some heat is left over in raw syngas Stream 5 and in the HRSG exhaust Stream 14. These are represented on Figure 5-5 by Trim 1 and Trim 2 respectively. Note that the Selexol solvent regeneration heat duty is not explicitly accounted for in the NETL report. However, Air Products' simulation suggests that there is ~288 MMBtu/hr of leftover heat between 315°F and 95°F available on the raw syngas. Air Products estimates that this heat is more than what is necessary to supply the regenerating heat duty for the Selexol unit. In addition, there is ~104 MMBtu/hr of leftover heat between 321°F and 270°F available on the HRSG flue gas. Moreover, analysis of the cooling curves in the simulated HRSG shows a non-optimum use of the heat available between 350°F and 720°F, suggesting that a better heat integration could use some higher-grade heat as well. Fuel preheat to 420°F (per the NETL report) requires ~94 MMBtu/hr that are not explicitly reported in the NETL study. It was therefore assumed that there is enough heat left over in the HRSG to provide the heat duty necessary for the fuel preheat.

New process cycle using Sour PSA technology

The experimental results obtained in Tasks 3 and 4 were used to simulate the performance of the PSA system using Air Products' proprietary simulation software, SIMPAC. The simulation results, in combination with Air Products' engineering expertise and significant experience in PSA design, were used to determine the size and number of PSA vessels. The feed to the PSA, flow, temperature, pressure and composition, were taken from the shifted syngas exiting the mercury removal bed of the S3B reference case. [2] The Sour PSA design consists of three trains of 10 beds with dimensions equivalent to commercially available units. At steady state, the Sour PSA allows 92.7% of H_2, 90.9% of N_2, 88.0% of CO and 90.8% of Ar to be recovered in the product stream, while 95.3% of CO_2 and more than 99.9% of H_2S, COS and water are rejected in the PSA tail gas. The product gas is sent as fuel directly to the gas turbine for power generation. The tail gas containing CO_2 and all the sulfur products needs further treatment for disposition of the sulfur and production of a CO_2 stream clean enough to meet sequestration specifications. The PSA cycle chosen for this case is the classic one (no rinse step) Air Products developed for a previous DOE-sponsored project. [1] Indeed, the cycle described in Task-4 for the production of hydrogen for power applications while demonstrating a higher H_2 recovery does not produce a CO_2 stream that readily matches specifications for sequestration.

The PSA tail gas treatment process relies on commercially available technologies. The tail gas from the Sour PSA is sent to an acid gas enrichment (AGE) process where H_2S is separated by a solvent-based absorption/regeneration process. This forms a sufficiently concentrated stream to be sent to a Claus unit to dispose of the sulfur in its elemental form. The H_2S-depleted stream contains mostly CO_2, H_2 and N_2 and is sent to a partial condensation unit for CO_2 purification. The rejected impurity stream contains H_2 and other non-condensables, including combustibles like CH_4 and CO, and is blended with the gas turbine fuel.

The simulation of the Sour PSA with acid gas enrichment and Claus plant in the coal gasification to power process is shown in the simplified process flow diagram (PFD) provided in Figure 5-6.

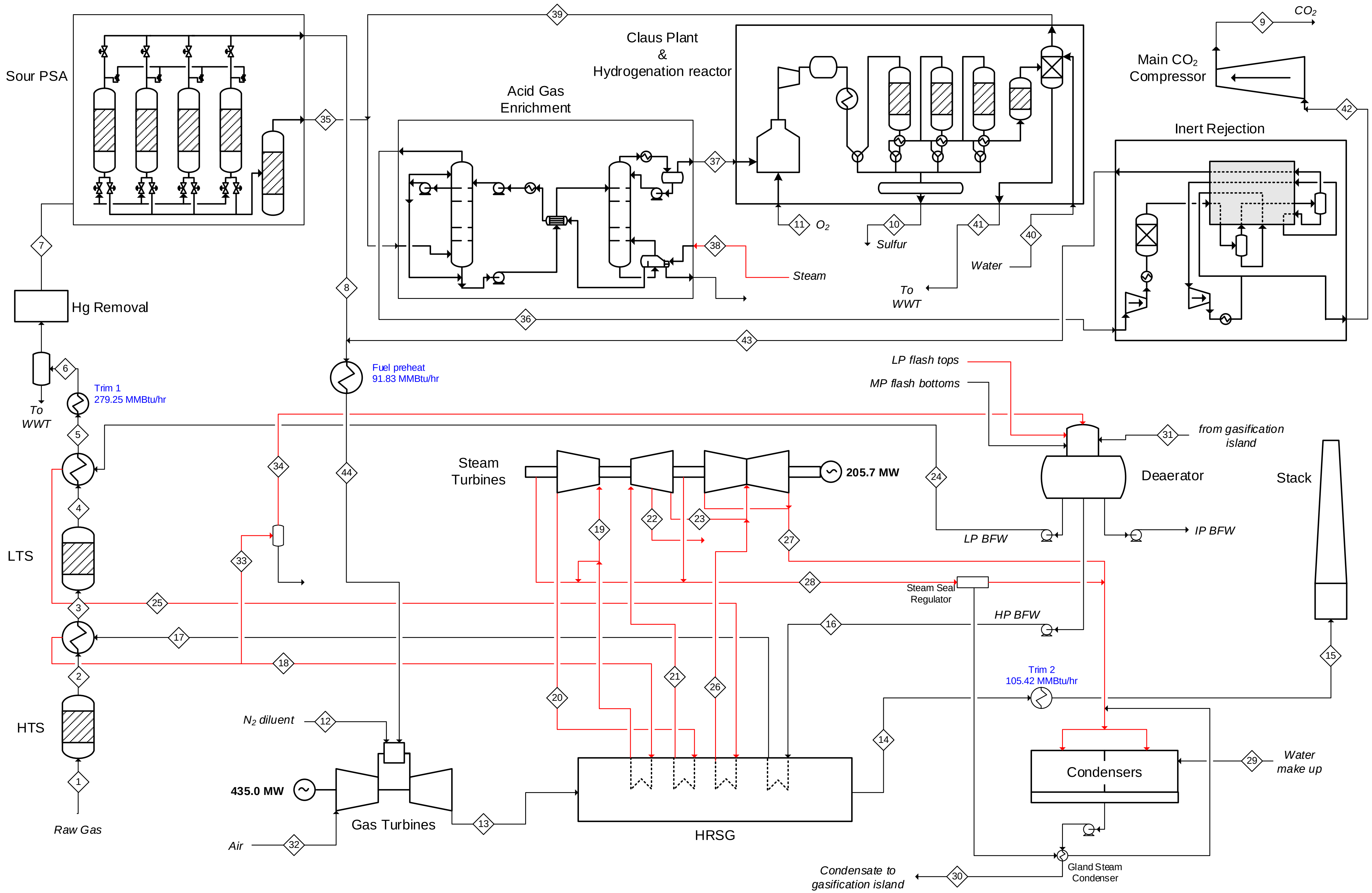


Figure 5-6: Simplified PFD of the simulated process with Sour PSA technology.

The AGE system performance was based on Exxon Mobil Flexsorb® technology and consists of an arrangement of absorber and stripper columns. The details of the process were not modeled; Air Products used results from a prior internal study and publically available information to extract correlations and calculate an overall heat and mass balance. Because the feed stream to the Claus unit has a similar composition to the one of the reference plant, the Claus plant is not modeled but instead scaled from the NETL S3B reference plant. [2]

The sweetened stream is sent to a light impurities rejection (partial condensation) unit that was originally developed by Air Products for oxyfuel CO_2 capture for pulverized coal combustion power boilers. The auto-refrigerated process involves a series of flash and cooling steps that require compression of the feed up to 522 psia. The process has a total auxiliary power load of 34.2 MW. Importantly, the CO_2 stream is available from the system at 240.5 psia, reducing the compression energy requirements downstream of the unit prior to entering the pipeline for end disposition. The CO_2 product contains 98.6% CO_2 and meets the NETL suggested specification except for the Ar level (446 ppm vs. <10 ppm) and N_2 level (4,081 ppm vs. <300 ppm). Further optimization of the process could be done to remove the impurities. However, the CO_2 purity criteria are not a strict requirement of the process and the optimization activity was not started. The overall carbon capture level achieved is 90.3% (92.6% CO_2 capture).

The rejected "lights" stream from the cold box is composed of 68.9% H_2, 19.8% CO_2, 8.8 % N_2, and 1.7% CO and has a Btu content of 194.2 Btu/SCF. The simulation takes advantage of this significant fuel value and the fact that this stream is already pressurized to blend it with the fuel for the gas turbines. The fuel for the gas turbines is obtained by using N_2 to dilute the Sour PSA product and the non-condensables from the light impurities rejection unit to achieve a fuel with a lower heating value (LHV) of 120 Btu/SCF (similar to the reference case). Consequently, there is ~1% more fuel available for the gas turbines than in the reference case, resulting in an equivalent increase of the gas turbines power output to 435 MW (vs 430.9 MW for the reference case) assuming constant heat rate.

The increase in the gas turbine power output also translates into more heat being available in the HRSG to produce more steam than in the reference case. Again, the approach taken here is to pass all the benefit of the increased heat available on raising more steam for the power-producing steam turbine, keeping the process steam production similar to the reference case. The additional steam is produced at the same conditions and HP, IP and LP ratios as in the reference case. Simulation of the steam cycle indicates that the additional available heat translates to an increase of 1% of the steam turbine power output to 205.7 MW (vs 203.8 MW for reference case). However, constraining the steam production to the same conditions as the reference case precludes full utilization of the additional heat available. Similar to the reference case (see above), a portion of the 105 MMBtu/hr (Trim 2) heat left over in the HRSG flue gas is enough to preheat the gas turbines fuel (~92 MMBtu/hr). However, as shown in Figure 5-6, there is still 279 MMBtu/hr available in the raw syngas (Trim1). The estimation of the steam needed for the solvent regeneration in the Selexol system in the base case is not readily attainable from the heat and material balance in the NETL report, but it was assumed that some of the heat from the raw syngas stream was used for the reboiler duty of the H_2S stripping column. Air Products estimated that only ~331 MMBTU/hr of LP steam is necessary for regenerating the Flexsorb® solvent in the stripper column of AGE unit of the Sour PSA process. This suggests that the leftover heat available in the raw syngas and HRSG is sufficient to generate enough steam for the duty of the reboiler to strip the H_2S in the regeneration column of the AGE system. A summary of the heat and mass balance for the simulated process is available in Appendix Table A5-2.

The choice of the AGE and Claus plant option for sulfur disposal in combination with the Sour PSA technology results in a small (1.0%) increase in the overall power production from both the gas and the steam turbines. This also leads to a small increase in the auxiliary power load, mainly due to a larger ASU and N_2 compressor required for the increased N_2 demand for fuel dilution, and a higher water circulation and cooling requirement due to the increased steam production. However, as can be seen in the power summary in Table 5-1, the use of the Sour PSA technology results in small increase of the net power produced.

Table 5-1. Summary of IGCC plant power output for base case and Sour PSA case.

POWER SUMARY (kWe)	Reference case	Sour PSA case
Gas Turbine	430,900	435,051
Steam Turbine	203,800	205,699
Total power	**634,700**	**640,750**
AUXILIARY LOAD (kWe)		
Coal Handling	510	510
Coal Milling	2,700	2,700
Slag Handling	580	580
WTA Coal Dryer Compressor	9,270	9,270
WTA Coal Dryer Auxiliary	600	600
ASU Auxiliary	1,000	1,000
ASU Main Air Compressor	62,000	62,768
O2 Compressor	8,670	8,666
N2 Compressor	34,640	35,069
CO2 Compressor	31,220	15,793
Boiler Feed Water Pumps	2,330	2,352
Condnesate Pump	220	222
Quench Water Pump	10	10
Circulating Water Pump	3,090	3,118
Ground Water Pump	360	363
Cooling Tower Fan	2,020	2,137
Air Cooled Condenser Fan	2,990	3,164
Scrubber Pumps	750	750
Acid Gas Removal/Sour PSA	18,190	-
AGE + inert removal	-	34,831
Gas Turbine Auxiliary	1,000	1,010
Steam Turbine Auxiliary	100	101
Claus Plant Auxiliary	250	239
Claus Plant TG Compressor	1,460	1,395
Misc. Balance of Plant	3,000	3,000
Transformer Losses	2,450	2,473
Total Auxiliary Power	**189,410**	**192,120**
NET POWER (kWe)	**445,290**	**448,629**

Economic analysis

Capital and O&M costs

Results from the process simulation as well as scaled equipment costs from the reference case S3B [2] were first updated to 2011 US$, then used to estimate the cost of the new process using Sour PSA technology. Air Products leveraged internal engineering resources to estimate the capital cost of new technology such as the Sour PSA, the inert rejection unit and the AGE equipment. Operating costs of the AGE (based on Flexsorb® technology) were estimated using internal work involving external partners as well as data published by Exxon.

Since the Sour PSA and the inert rejection unit are new technologies that have not been implemented at commercial scale for IGCC applications, a 20% contingency was used for both the process and the project. This is certainly a conservative approach for the Sour PSA systems given Air Products' extensive experience with H_2 PSA units and cold box technologies at commercial scale. It would be reasonable to expect that one of the contingencies, either process or project, would be reduced due to the commercial readiness of the individual steps involved.

The Total Plant Cost (TPC) of the Sour PSA case compares favorably with the reference case (5.7% cheaper), as seen in Table 5-2. The TPC details for both the reference plant and the Sour PSA case are available in Appendix Tables A5-3 and A5-4 respectively. The main difference is that the acid gas removal and CO_2 compression total cost is ~44% lower (~$105MM) for the Sour PSA technology.

Table 5-2. High-level summary of IGCC TPC for base case and Sour PSA case.

	Reference case	Sour PSA case
Total plant cost ($x1000)	1,753,599	1,653,205
Total plant cost ($/kW)	3,938	3,685

Both fixed and variable operating costs have been estimated on the same basis that NETL used for the reference case. [2] The cost of the Sour PSA adsorbent is provided for both the initial load and as a yearly cost, although it is expected that the adsorbent will be replaced only every 10 years.

Both the reference plant and the Sour PSA case are located in Montana, which results in specific price for fuel and CO_2 Transport, Sequestration and Monitoring (TS&M). Following guidelines from the "Fuel price for selected feedstock" NETL report [7] a cost of coal of $19.63 per ton was used. In addition, Air Products used a TS&M cost of $22 per metric tone of CO_2 according to the "Updated cost for selected bituminous baseline case" NETL report [3] to estimate the annual cost of CO_2 TS&M. Note that the "Cost and performance baseline for fossil energy plants" 2008 report [2], from which the IGCC reference case was developed, did not include the CO_2 TS&M cost in the operating cost calculation. Rather, it used a $6.30 per MWe additional cost in the Cost of Electricity (COE) calculation, which corresponds to about $6.49 per ton of CO_2. For the present study, it was decided to include the CO_2 TS&M cost in the variable operating cost for both the reference and the Sour PSA cases.

Table 5-3 provides a high-level summary of the annual Operating and Maintenance (O&M) cost for both the reference and the Sour PSA cases.

Table 5-3. High-level summary of IGCC O&M cost for base case and Sour PSA case.

	Reference case	Sour PSA case
Net Power output (kW)	445,290	448,629
Fixed operating cost ($/year)	64,151,703	60,996,683
Variable operating cost $/year)	106,349,791	105,209,600
Fuel cost ($/year)	39,882,511	39,882,511

The details of the calculation of the O&M cost is provided in the Appendix in Tables A5-5 and A5-6 for the reference case and the Sour PSA case. Using the Sour PSA rather than the incumbent technology brings significant savings on the operating cost, primarily by driving down fixed operating costs by ~5%. This is mainly the result of lower maintenance and insurance costs due to lower capital cost.

Financial analysis

Cost of electricity

In the “Cost and performance baseline for fossil energy plants” NETL report [2], the Cost Of Electricity (COE) is calculated using the Total Overnight Cost (TOC) and a Capital Charge Factor (CCF) rather than from a cash flow analysis. In a previous DOE-sponsored study [1], Air Products showed that there were minor differences in the COE depending on the calculation method used, and that the way the CO_2 TS&M was accounted for was a major contributor in the difference. Nevertheless, since goal was to compare the new Sour PSA technology against an incumbent one, the method used was irrelevant as long as both the reference and the new cases were compared using the same methodology.

The Statement of Project Objectives (SOPO) for the present study suggests using the CFF method, with the cash flow analysis being optional. It was decided to provide both approaches for the IGCC cases. The main reason for providing the cash flow analysis for IGCC is that it was the only method that could also be used for the methanol cases, since Air Products was unable to reconcile the SOPO guidance for CCF and the methodology used in the “High value gasification products” NETL report. [8]

In order to establish the TOC for both cases, the owner’s cost was first estimated according to the methodology developed in the “Cost estimation methodology” NETL report. [4] Table 5-4 provides a summary of the owner’s cost for both the reference and the Sour PSA cases. The owner’s cost is ~5.6% lower for the Sour PSA case, again driven mostly by the lower TPC than in the reference case.

Table 5-4. Estimation of owner's cost for base case and Sour PSA case.

Item		Reference case	Sour PSA case
Start up cost		**54,791**	**51,894**
	6 months operating labor	14,540	13,966
	1 month maintenance materials at full capacity	3,190	3,007
	1 month non fuel consumable at full capacity	500	366
	1 month waste disposal	452	452
	25% of one month's fuel cost at full capacity	1,039	1,039
	2% of TPC	35,072	33,064
Inventory Capital		**17,948**	**17,478**
	0.5% of TPC	8,768	8,266
	60 days of supply (full capacity) of fuel (n/a for NG)	8,195	8,195
	60 days of supply (full capacity) of non-fuel consumables.	985	1,017
Land ($x1000) ($3000/acre)		**900**	**900**
Financing cost	(2.7% of TPC)	**47,347**	**44,637**
Other Owner Cost	(15% of TPC)	**263,04**	**247,981**
Initial Catalyst and Chemical Cost		**15,245**	**13,944**
Total Owner's Cost ($x1000)		**399,272**	**376,833**

The TPC and the owner's cost were used to calculate the TOC as well as the Total As Spent Cost (TASC) using a TASC multiplier of 1.14 corresponding to the financial structure for a high-risk Investor Owned Utilities (IOU) project. [4] Results are shown in Table 5-5. Note that the TASC is not directly used in the calculation of the cost of electricity, which rather relies on TOC and an ad hoc CCF.

Table 5-5. Estimation of TOC and TASC for base case and Sour PSA case.

	Reference case		Sour PSA case	
	$x1000	$/kW	$x1000	$/kW
Total Plant Cost	1,753,599	3,938	1,653,205	3,685
Owner's Cost	399,272	897	376,833	840
Total Overnight Cost	2,152,870	4,835	2,030,038	4,525
Total As Spent Cost (x1.14)	2,454,272	5,512	2,314,244	5,158

The TOC and O&M cost are then used to estimate the cost of electricity (COE) using a 0.124 Capital Charge Factor (CCF) that represents the financial structure for a high-risk Investor Owned Utilities (IOU) project with a 5-year capital expenditure period. [4] Table 5-6 provides a comparative breakdown of the COE for both the reference and the Sour PSA cases.

Table 5-6. Estimation of COE for base case and Sour PSA case using the CCF method.

	Reference case	Sour PSA case
Total Overnight Cost ($)	2,152,870,284	2,030,038,393
Fixed Operating Cost FOC ($/year)	64,151,703	60,996,683
Variable Operating Cost VOC ($/year)	146,232,301,	145,092,111
Production (MWh/year) @ 80% CF	3,122,730	3,146,148
Cost of Electricity ($/MWh)	**153.07**	**145.71**
Fuel ($/MWh)	12.77	12.68
FOC ($/MWh)	20.54	19.39
VOC ($/MWh)	12.72	12.12
Capital ($/MWh)	85.69	80.20
CO2 TS&M ($/MWh)	21.33	21.32

The combination of lower capital cost, lower fixed and variable costs and slightly higher electricity production (thanks to lower parasitic power) results in a 4.8 % lower cost of electricity for the Sour PSA case compared to the case using incumbent technology.

While quick and certainly valid for case comparisons, the above method does not provide the accuracy of a cash flow analysis to determine the COE. A cash flow analysis was performed using the global economic assumptions (Table 5-7) established in the "Cost estimation methodology for NETL assessment of power plant performance" NETL report. [4]

Table 5-7: Global economic assumptions.

Item	Description
Taxes	
Income Tax Rate	38% (Effective 34% Federal, 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	0%
Tax Holiday	0 year
Contracting and Financing Terms	
Contracting Strategy	EPC managemnt (owner assumes project risk)
Type of Dept Financing	Non Recourse
Repayment term of Debt	15 years
Grace Period on Dept Repayment	0 year
Debt Reserve Fund	none
Analysis Time Periods	
Capital Expenditure Period	5 years
Operational Period	30 years
Economic Analysis Period (for IRROE)	35 years
Treatment of Capital Costs	
Cap. Cost Escalation during CAPEX period	3.6% (average from Chemical Engineering Plant Cost Index)
Distribution of TOC over CAPEX period	5 years: 10%, 30%, 25%, 20%, 15%
Working Capital	0
% of TOC being Depreciated	100%
Escalation of Operating Revenues and Cost	3% from average DoL Producer Price Index for Finished Goods

This analysis also used the financial structure for a high-risk IOU suggested in the "Recommended project finance structure" NETL report. [5] It assumed a 5-year capital expenditure for the project. Financing was a) 45% through debt at a 3.5% LIBOR rate majored by 2%, and b) 55% through equity at an expected return of 12%. These assumptions are summarized in Table 5-8.

Table 5-8. Financial structure for high-risk IOU projects.

	% of Total	Current $ cost	Current weighted cost	After tax weighted cost
Debt	45	5.5%	2.475%	
Equity	55	12%	6.60%	
Total			9.075%	8.13%

The cash flow analysis was used to determine the COE assuming a 5-year capital expenditure, 35-year analysis period, 15-year debt repayment period, 20-year depreciation period, and a 12% ROE for the project. The details for both the reference and Sour PSA cases are presented in Appendix Tables A5-7 and -8 respectively. Table 5-9 summarizes the comparison of both cases.

Table 5-9. COE and project NPV results from the cash flow analysis for both the reference and the Sour PSA cases.

	Reference case	Sour PSA case
Cost of electricity ($/MW)	157.01	149.78
Fuel ($/MWh)	12.77	12.68
FOC ($/MWh)	20.54	19.39
VOC ($/MWh)	12.72	12.12
Capital ($/MWh)	89.65	84.27
CO_2 TS&M ($/MWh)	21.33	21.32
NPV at discount rate ($x1000)		
8%	581,969	548,668
10%	231,286	218,051
12%	0	0

Similar to what Air Products found in a previous DOE study [1], the COE calculated by cash flow analysis is slightly higher than that determined using the CCF method. More importantly, the significant economic advantage of Sour PSA over the incumbent technology is still demonstrated by a 4.6% lower COE. These values will subsequently be used in this report as the estimated COE.

Sensitivity of the COE

The cash flow analysis was used to perform sensitivity analysis around the capital cost of the Sour PSA technology, selling price of CO_2 at the plant's gate, and cost of CO_2 emissions. The COE sensitivity to the capital cost of the Sour PSA technology was established by varying the total cost of the new technologies for the AGR scope (Sour PSA, AGE, inert removal) from 50% to 200% of the value used to establish the COE in Table 5-9. The resulting evolution of the COE is presented in Figure 5-7.

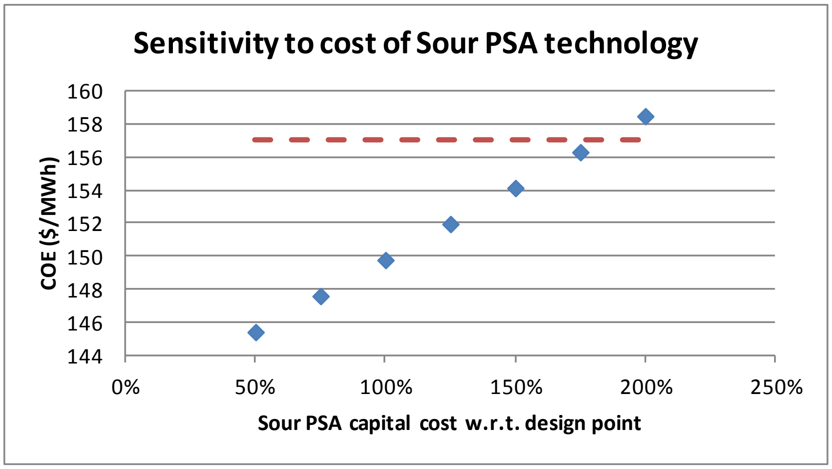


Figure 5-7. Evolution of the COE with the capital cost of the Sour PSA AGR technology.

The doted red line corresponds to the COE for the reference case. This suggests that the capital cost of all the equipment used for AGR with the Sour PSA technology would have to be more than 83% higher than our assumptions to result in a higher COE than the reference technology can produce.

The sensitivity of the COE to the performance of an AGR system based on the Sour PSA technology is more complex to establish. Indeed, lower performance of an AGR system means that products and/or recovery are not at the designed specifications. This can always be mitigated with larger-capacity or additional equipment and therefore would fall into the capital cost sensitivity presented above. Rather, Air Products decided to look at the sensitivity of the COE to the parasitic load of the AGR system since it is also a performance factor. Therefore, the evolution of the COE was assessed when varying the power consumption of the entire AGR system (Sour PSA, AGE, inert removal) from 50% - 200% of the design point used to establish the COE in Table 5-9. The resulting evolution of the COE is presented in Figure 5-8.

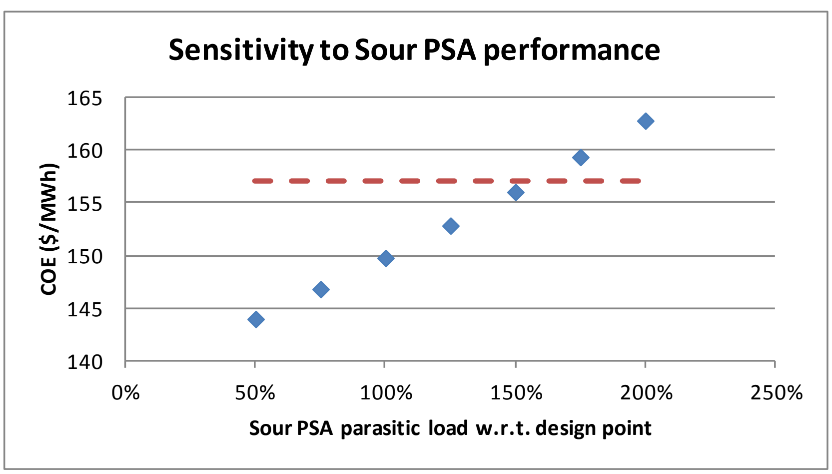


Figure 5-8. Evolution of the COE with power load of the Sour PSA AGR technology.

The doted red line corresponds to the COE for the reference case. This suggests that the parasitic power for Sour PSA-based AGR would have to be more than 56% higher for this technology to be less economical than the reference case.

The COE evolution was compared with the selling price of CO_2 for both the reference and Sour PSA cases. In this scenario, all the captured CO_2 is sold at the gate and no TS&M is incurred. The CO_2 price is established for the first year of operation (0-60 $/tonne) and increased by 3% every year according to the global economic assumptions in Table 5-7. The results suggest that the CO_2 would have to be sold ~18.5% higher for the reference case to be as economical as the Sour PSA case.

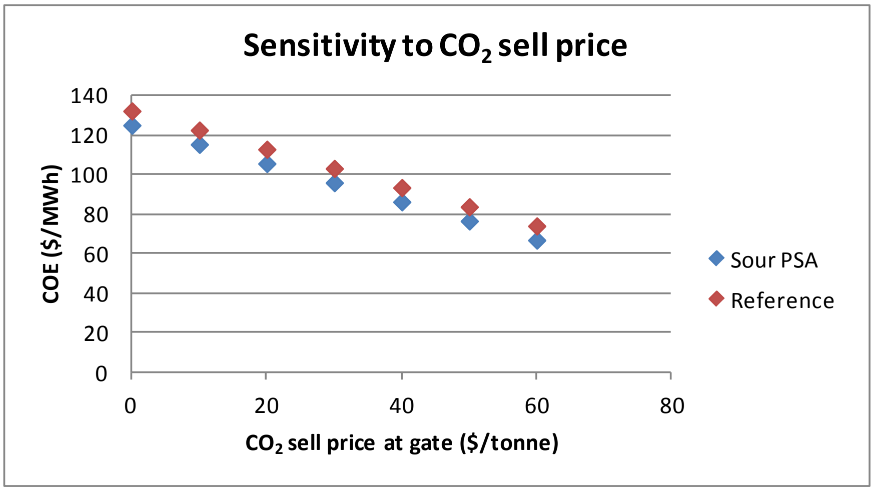


Figure 5-9. Comparison of the evolution of the COE with sale price of CO_2.

Finally, the sensitivity of the COE was assessed relative to a tax on the emitted CO_2. In this scenario, the tax is paid only on the CO_2 that is not captured by the AGR technology (exhaust of the gas turbine). Again, the reference and Sour PSA cases were compared. The tax is fixed at the first year of operation (0-$60/tonne) and increased by 3% every year according to the global economic assumptions in Table 5-7. The resulting evolution of the COE is presented in Figure 5-10.

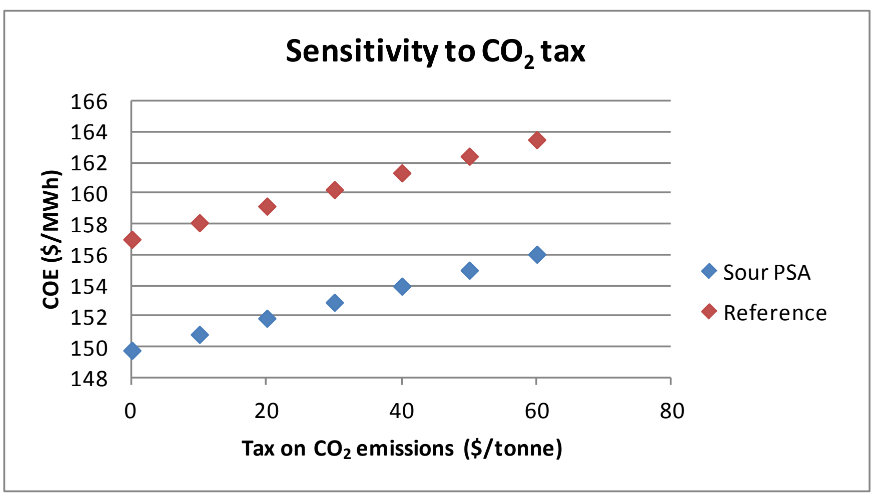


Figure 5-10. Comparison of the evolution of the COE with tax level on CO_2 emissions.

Again, the results show that regardless of the tax on the emitted CO_2, the Sour PSA-based AGR always results in a lower COE than the incumbent AGR. The higher COE of the reference case (without CO_2 tax) is equivalent to a tax of $69.25/tone of CO_2 on the emissions of the Sour PSA case.

Techno-economic evaluation of Sour PSA technology for coal-to-methanol applications

A techno-economic assessment was also completed for the impact on the Cost Of Methanol (COM) from incorporating Air Products' Sour PSA technology into a 50,000 barrel per day coal-to-methanol plant with CO_2 capture.

The methodology used here borrows from both the 2011 "Cost performance baseline for fossil energy plants" report by NETL [2] and the more methanol-specific 2014 "High Value Gasification Products: Crude Methanol Cases"[8] NETL reports. This base case uses the same technology as described in Case 2 of the second report [8], but has been scaled down to produce 50,000 barrel per day of methanol. The reference system (Figure 5-11) consists of a coal handling and drying system that feeds trains of Shell® gasifiers and an air separation unit (ASU) providing O_2 to the gasifier. The raw syngas produced by the gasifier is scrubbed and then partially shifted to H_2 and CO_2 prior undergoing Hg removal and being sent to the AGR unit. The AGR consists of a Rectisol® unit and a Claus plant with tail gas recycled for coal drying. The sweet syngas is sent to the methanol plant and the CO_2 product is compressed to pipeline specifications. Power needs are met though a NGCC plant with an Econamine® CO_2 removal unit.

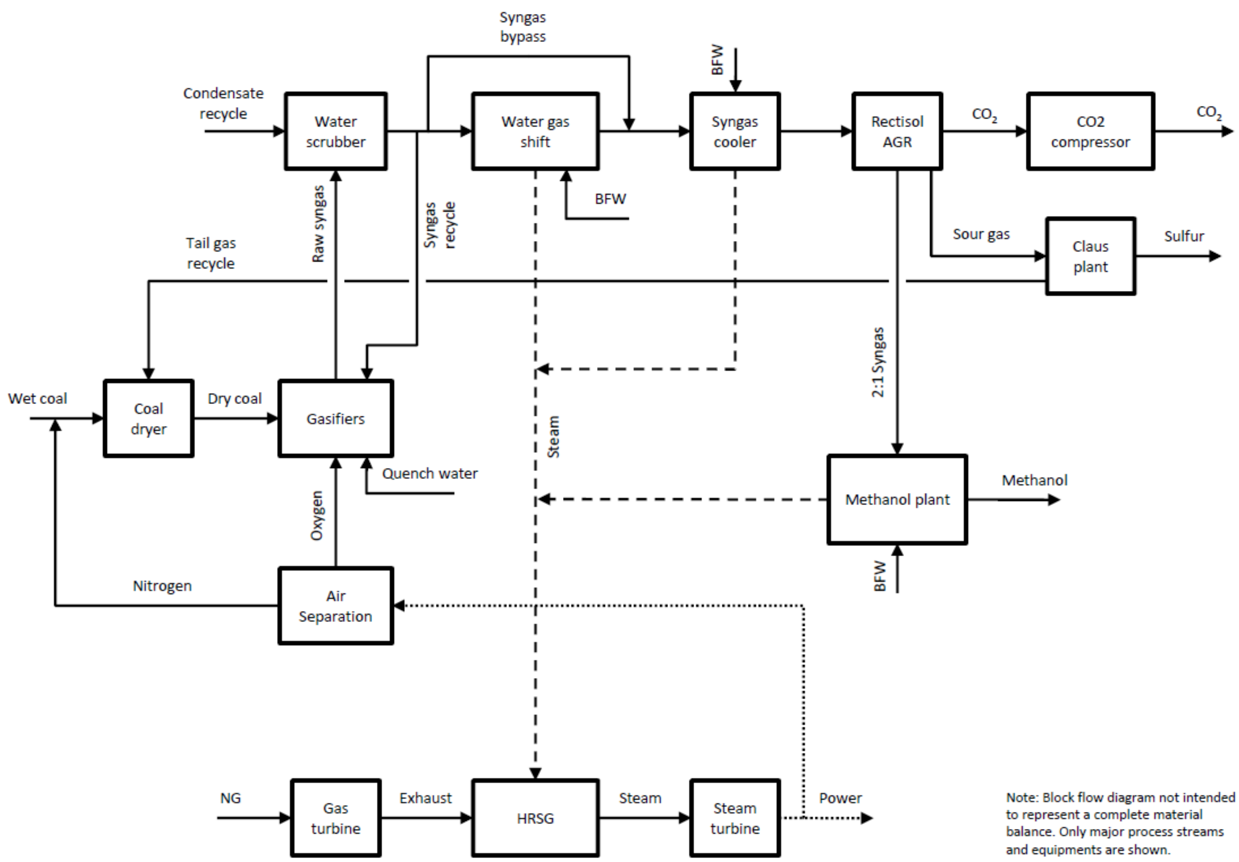


Figure 5-11. Basic bloc flow diagram of the reference coal-to-methanol plant.

To enable a comparison between the Sour PSA and reference case, the gasification trains were kept at the same size. The gas turbine was kept the same, but because of different heat integration the HRSG and steam turbine were scaled down for the Sour PSA case.

No process simulation was done for the reference case since it was assumed that the performance would linearly scale with the coal feed rate. Therefore, the data used was scaled from the "High Value Gasification Products" NETL report. [8] For the Sour PSA-based AGR case, the shift reactors, Sour PSA and AGE were simulated in steady state at a high level using Aspen Plus.®

Process selection

Both the reference case and the plant using the Sour PSA technology were assumed to be located in the midwest and use delivered PRB coal as feedstock. The cost of fuels and cost of CO_2 Transport and Storage (T&S) were calculated according to NETL guidelines. [7, 3]

Reference case

The reference plant used in this analysis is a scaled-down version of Case 2 described in the "High Value Gasification Products: Crude Methanol Cases" [8] NETL report. The plant's coal feed rate was scaled down (x 0.64) so that the output would be ~50,000 barrel per day (bpd) of crude methanol. Such a plant would consist of:

- Two ASU trains
- Two train of coal drying and feed system
- Four trains of gasification including gasifier (Shell), syngas cooler, particulate removal
- Two trains of syngas cleanup process
- One train of Rectisol® acid gas removal system
- One train of Claus based sulfur recovery
- One methanol plant
- One NGCC plant with Econamine® CO_2 capture technology

New case using Sour PSA Technology

The results from the experimental tests conducted in Tasks 3 and 4 provided a basis on which to model, design, and cost a process configuration for the TEA. The process design for the new case is identical to the reference case except that the Rectisol® unit is replaced by a Sour PSA based acid gas removal system. Like in the IGCC configuration, the tail gas from the Sour PSA is treated with a Flexsorb® AGE unit so that sulfur can be recovered through a Claus plant.

Product recovery for H_2 and CO with Sour PSA alone is not 100% since some of it ends up in the tail gas. In the IGCC configuration, a partial condensation unit is used to recover these lights elements from the CO_2 stream after the AGE step, resulting in a 99.5% overall H_2 recovery. Since the CO_2 stream is meant to be compressed for sequestration anyway, the use of a partial condensation unit based on auto-refrigeration leads to only a very small parasitic power penalty but adds some capital. Air Products has also developed new PSA cycles including an intermediate rinse step using the PSA tail gas (see Task 4) to significantly improve product recovery with very limited capital expenditure. Internal Air Products studies have shown that these new cycles provide significant advantage in non-CO_2-capture applications. Therefore, the Sour PSA case was designed for methanol production using the PSA rinse cycle instead of the partial condensation unit to optimize product recovery.

Process simulation

As mentioned earlier, the reference plant process was not simulated directly. Instead, Air Products used the heat and mass balance provided in the "High Value Gasification Products" NETL report [8] and scaled down the coal feed rate so that the overall methanol production was about ~50,000 bpd. The assumption was that plant performance would remain the same and that gasifier and downstream equipment would scale according to the coal feed rate. The process duty for the Rectisol® unit was estimated from a combination of publicly available literature [10, 11] and previous Air Products engineering studies. For the Sour PSA case, only the AGR process and the part of the process that would be impacted by the use of Sour PSA were simulated in Aspen Plus.® The performance of Sour PSA using the new PSA cycles was estimated using Air Products' proprietary SIMPAC simulation tool.

Reference process cycle

The reference cycle used is the one described in the "High Value Gasification Products" NETL report [8], except that the coal feed rate was scaled down from 1,618,190 lb/hr to 1,041,740 lb/h (i.e., a 0.64 multiplying factor). All downstream material flows were therefore scaled down by the same factor. However, the utility requirements for the Rectisol® unit process are not entirely specified in the NETL report [8] from which this base case is derived. Based on the amount of absorbed gas, the power requirement was estimated to be ~36.9 MW (solvent circulation pumps and refrigeration), and the steam requirement to be 163.6 MMBTU/h (72 PSI saturated steam) for the Rectisol® unit for this base case. Figure 5-12 shows a simplified version of the PFD for the reference cycle. The heat and mass balance summary for the major material stream is available in Appendix Table A5-9.

In addition, the heat duty of the HRSG from the power plant had to be scaled differently because this case uses the same gas turbine and Econamine® system as the reference, but the heat recovered from the rest of the plant had to be scaled down (Table 5-10). A pinch analysis showed that the reduced amount of steam produced by the HRSG would result in only 193.5 MW produced by the steam turbines.

Table 5-10. High-level summary of the HRSG heat duty.

	Heat duty (MMBtu/hr)
Recovered from integration	1,140.7
Recovered from gas turbine	531.8
Needed for Econamine® unit	202.5
Available for steam turbine	1,470

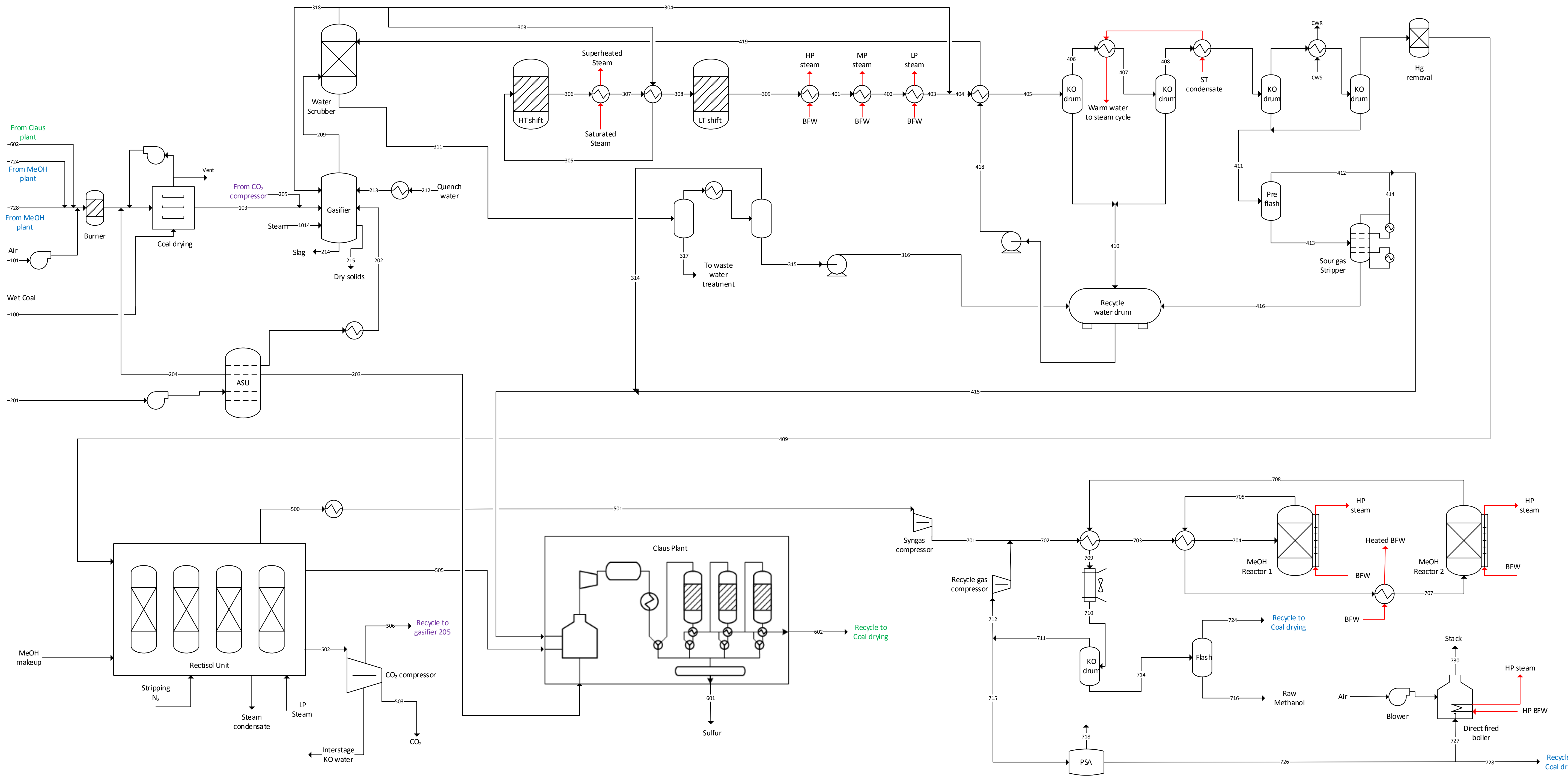


Figure 5-12. Simplified PFD of the coal-to-methanol reference cycle.

New process cycle using Sour PSA technology

The experimental results obtained in Task 3 and 4 were used to estimate the performance of the Sour PSA technology using Air Products' SIMPAC proprietary simulation system. A newly developed PSA cycle was used, which included a rinse step to improve the overall product recovery without having to add equipment to recover product from the tail gas. In this cycle, ~18% of the PSA tail gas is recompressed and used as a rinse stream during the first equalization step of the cycle. In order to produce a sweet syngas suitable for methanol production (2:1 H_2 to CO ratio and Syngas Number SN~2), Air Products optimized the bypass ratio for the shift reactor. Both the Sour PSA and the shift reactors were simulated in Aspen Plus,® while the performance of the Sour PSA was estimated with SIMPAC. The optimum solution was obtained with 40 % of the syngas bypassing the shift reactors. The simulation results were used along with Air Products' expertise in engineering and PSA design to estimate the size and number of PSA vessels needed to treat enough syngas to produce 50,000 bpd of methanol. Under these conditions, the Sour PSA design consists of six trains of 10 beds with dimensions equivalent to available commercial units. At steady state, the Sour PSA allows 96.7% of H_2, 96.4% of N_2, 93.3% of CO and 96.4% of Ar from the feed to be recovered in the product stream, while 98.2% of CO_2, and >99.99% of H_2S, COS and water are rejected in the PSA tail gas. The power requirement for the Sour PSA tail gas compressor used for the rinse step was estimated to be 7.8 MW.

The tail gas containing CO_2 and all the sulfur products needs further treatment for disposition of the sulfur and production of a CO_2 stream clean enough to meet sequestration specifications. Like for the IGCC case, the selected PSA tail gas treatment process relies on commercially available technologies. The tail gas from the Sour PSA is sent to an acid gas enrichment (AGE) process where H_2S is separated by a solvent-based absorption/regeneration process. This forms a sufficiently concentrated stream to be sent to a Claus unit to dispose of the sulfur in its elemental form. The H_2S-depleted stream contains mostly CO_2 which is ready for sequestration. The implementation of the Sour PSA with AGE and a Claus plant in the coal gasification-to-power process that was simulated is described on the simplified PFD provided in Figure 5-13.

The AGE system performance was based on Exxon Mobil Flexsorb® technology and that consists of an arrangement of absorber and stripper columns. The model does not contain the full details of the process; it uses results from a prior internal study and publically available information to extract correlations and calculate an overall heat and mass balance. It was estimated that ~450 MMBtu/hr of LP steam are necessary for regenerating the Flexsorb® solvent in the stripper column of AGE unit. The models suggested a H_2S content of 27.6% in the enriched stream to be sent to the Claus plant. Because the feed stream to the Claus unit has a composition similar to that in the reference plant, the Claus plant was not modeled but instead scaled from the reference plant case. Overall, 85.8% of the CO_2 present in the syngas entering the AGR is captured compared to 76.5% for the reference case.

In order to account for the much lower syngas CO_2:CO ratio than for the reference case, Air Products also simulated the methanol plant process. Overall, the methanol production was reduced by ~2% when using the Sour PSA technology. The heat and mass balance summary for the major material stream is available in Table A5-10 in the appendix.

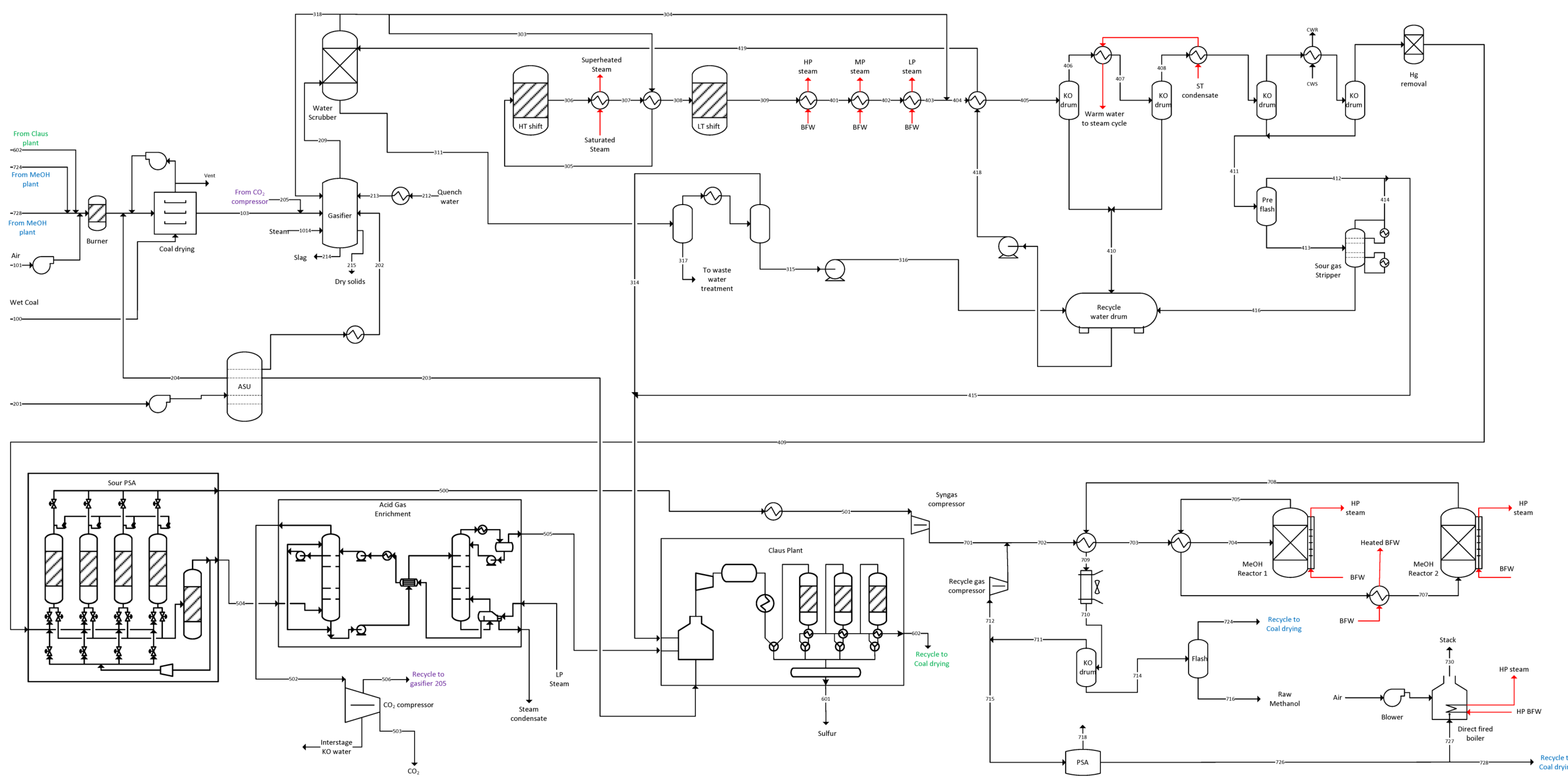


Figure 5-13. Simplified PFD of the coal-to-methanol cycle with Sour PSA.

The auxiliary power plant was considered to be identical to the reference case with heat integration between the NGCC's HRSG and the rest of the plant. However, because of the higher steam requirement for the AGE system than was estimated for the reference Rectisol® unit, the steam turbine output is reduced to 155.7 MW (vs 193.5 MW) for the Sour PSA case.

The auxiliary power load is ~26.3 MW lower for the Sour PSA case compared to the reference case. Overall, the Sour PSA technology results in ~11.4 MW less power being available for export when compared to the reference case. Table 5-11 shows the power summary for both the reference and the Sour PSA cases.

Table 5-11. Summary of auxiliary plant power output for base case and Sour PSA case.

POWER SUMARY (kW)	Reference case	Sour PSA case
Gas Turbine	113,700	113,700
Steam Turbine	193,480	155,751
Total power	**307,180**	**269,451**
AUXILIARY LOAD (kW)		
Coal handling	5,852	5,852
Slag handling	1,249	1,249
ASU	115,840	115,840
Syngas Recycle Compressor	4,249	4,249
Incinerator air blower	1,725	1,725
Direct fired boiler blower	200	186
Flash bottoms pump	464	464
Scrubber Pumps	689	689
AGR auxiliary	36,895	338
PSA Tailgas Compressor	-	7,838
Claus Plant Auxiliary	161	155
CO2 compressor	46,319	50,135
Syngas compressor	13,365	12,426
Recycle gas compresoor	2,170	2,017
Water treatment	2,273	2,273
Air cooler fans	1,177	1,177
Circulating water pumps	6,071	6,071
Boiler feedwater pump	966	966
Cooling tower fans	333	333
Steam Turbine Auxiliary	70	56
Misc. Balance of Plant	3,273	2,969
NGCC plant	21,480	21,480
Total Auxiliary Power	**264,818**	**238,487**
NET POWER (kW)	**42,362**	**30,965**

Economic analysis

Capital and O&M costs

Results from the process simulation and scaled equipment cost (mainly from the "High value gasification products" NETL report [8]) were used to estimate the cost of both the base case plant and the new process using Sour PSA technology.

As with the IGCC study, Air Products leveraged internal engineering resources to estimate the capital cost of new technology such as the Sour PSA, inert rejection unit and AGE equipment. The operating cost of the AGE (based on Flexsorb® technology) was estimated using internal work involving external partners as well as data published by Exxon. Operating and maintenance costs for the reference case were scaled from the NETL report [8].

The same conservative approach was taken as with the IGCC case, using a 20% contingency for both project and process for the Sour PSA technology. Indeed, despite Air Products' extensive experience in PSA, this technology has never been implemented at that scale in coal-to-methanol applications. It would be reasonable to expect that the contingencies, either process or project, would be reduced due to the commercial readiness of the individual steps involved.

Like for IGCC applications, the Total Plant Cost of the Sour PSA case compares favorably with the reference case (8.7 % cheaper), as seen in Table 5-12. The details of the Total Plant Cost (TPC) for both the reference plant and the Sour PSA case are available in Appendix Tables A5-11 and A5-12 respectively. The main difference is that the total cost for acid gas removal and CO_2 compression is about ~39% (~$279MM) lower for the Sour PSA technology.

Table 5-12. High-level summary of TPC for base case and Sour PSA case.

	Reference case	Sour PSA case
Total plant cost ($x1000)	3,597,216	3,282,170
Total plant cost ($/bpd)	70.37	65.54

Both fixed and variable operating costs have been estimated on the same basis that NETL used for the reference case. [8] The cost of the Sour PSA adsorbent is provided for both the initial load and as a yearly cost, although it is expected that the adsorbent will be replaced only every 10 years.

Both the reference plant and the Sour PSA case are located in midwestern USA, which results in a specific price for fuel and CO_2 Transport, Sequestration and Monitoring (TS&M). Following guidelines from the "Fuel price for selected feedstock" NETL report [7], this study used a cost of $36.57 per ton for coal and $6.13/ MMBtu for natural gas. this study also used a TS&M cost of $11 per metric tonne of CO_2 according to the "Updated cost for selected bituminous baseline case" NETL report [3] in order to estimate the annual cost of CO_2 TS&M. Both cases accounted for revenues from selling power at $59.59/MW.h

Table 5-13 provides a high-level summary of the annual Operating and Maintenance (O&M) cost for both the reference and the Sour PSA cases. Following the methodology used in the "High value gasification products" NETL report [8], the revenues of the sale of electricity were included as a reduction of the variable operating costs. Details of the O&M cost calculations are provided as Appendix Tables A5-13 and A5-14 for the reference case and the Sour PSA case respectively.

Table 5-13. High-level summary of IGCC O&M cost for base case and Sour PSA case.

	Reference case	Sour PSA case
Net MeOH production (bpd)	51,119	50,079
Fixed operating cost ($/year)	125,445,107	115,349,469
Variable operating cost ($/year)	87,480,108	96,996,248
Fuel cost ($/year)	207,435,799	207,435,799

Using the Sour PSA technology for acid gas removal brings a significant fixed operating cost reduction, mainly driven by the much lower capital compared to the incumbent technology. However, the variable operating cost is ~9.8% lower for the incumbent technology. The main drivers for the higher variable operating cost for the Sour PSA technology are lower electricity sales and more CO_2 being sent for sequestration. It is important to note that these are the results of design trade-offs and are not specific to the technology.

Financial analysis

Cost of methanol

The authors of "High value gasification products" NETL report [8] calculated the Required Sale Price (RSP) of methanol using the Total Overnight Cost (TOC), the Total As Spent Cost (TASC) and a Capital Charge Factor (CCF) rather than from a cash flow analysis. While it is not clear what CCF was used, it does not match the 0.237 value Statement Of Project Objectives for this project. Therefore, for consistency Air Products calculated the TOC and TASC but used a cash flow analysis to calculate the methanol RSP.

To establish the TOC for both cases, the owner's cost was first estimated according to the methodology developed in the "Cost estimation methodology" NETL report. [4] Table 5-14 provides a summary of the owner's cost for both the reference and the Sour PSA cases. The owner's cost is ~6.2% lower for the Sour PSA case, again driven mostly by the lower TPC compared to the reference case.

Table 5-14. Estimation of owner's cost for base case and Sour PSA case.

Item		Reference case	Sour PSA case
Start-up cost		**110,073**	**101,673**
	6 months operating labor	26,750	24,853
	1 month maintenance materials at full capacity	4,818	4,396
	1 month non fuel consumable at full capacity	730	947
	1 month waste disposal	1,029	1,032
	25% of one month's fuel cost at full capacity	4,802	4,802
	2% of TPC	71,944	65,643
Inventory Capital		**46,854**	**45,707**
	0.5% of TPC	17,986	16,411
	60 days of supply (full capacity) of fuel (n/a for NG)	27,429	27,429
	60 days of supply (full capacity) of non-fuel consumables.	1,438	1,876
Land ($x1000) ($3000/acre)		**900**	**900**
Financing cost	(2.7% of TPC)	**97,125**	**88,619**
Other Owner Cost	(15% of TPC)	**539,582**	**492,326**
Initial Catalyst and Chemical Cost		**5,960**	**21,970**
Total Owner's Cost ($x1000)		**800,494**	**751,194**

The TPC and the owner's cost were then used to calculate the TOC and the Total As Spent Cost (TASC), as shown in Table 5-15, using a TASC multiplier of 1.181 that was back-calculated from the NETL report [8] for commercial fuel finance structure cases.

Table 5-15. Estimation of TOC and TASC for base case and Sour PSA case.

	Reference case		Sour PSA case	
	$x1000	$/bpd	$x1000	$/bpd
Total Plant Cost	3,597,216	70.4	3,282,170	65.5
Owner's Cost	800,494	15.7	751,194	15.0
Total Overnight Cost	4,397,710	86.0	4,033.364	80.5
Total As Spent Cost	5,193,695	101.6	4,763,403	95.1

A cash flow analysis was used to estimate the RSP for methanol in a more rigorous way than would be possible if using the TOC and a capital charge factor. This analysis used the same global economic and financial assumptions as in the "High value gasification products" NETL report. [8] Economic assumptions are listed in Table 5-16 and, except for the longer (30-year) debt repayment period, are identical to those used for the IGCC cases. The financial assumptions for commercial fuel projects are summarized in Table 5-17. Air Products assumed a 5-year capital expenditure for the project. Financing was 50% through debt at a 3.5% LIBOR rate majored by 4.5%, and 50% through equity at an expected return of 20%.

Table 5-16. Global economic assumptions for coal-to-methanol applications.

Item	Description
Taxes	
Income Tax Rate	38% (Effective 34% Federal, 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	0%
Tax Holiday	0 year
Contracting and Financing Terms	
Contracting Strategy	EPC managemnt (owner assumes project risk)
Type of Dept Financing	Non Recourse
Repayment term of Debt	30 years
Grace Period on Dept Repayment	0 year
Debt Reserve Fund	none
Analysis Time Periods	
Capital Expenditure Period	5 years
Operational Period	30 years
Economic Analysis Period (for IRROE)	35 years
Treatment of Capital Costs	
Cap. Cost Escalation during CAPEX period	3.6% (average from Chemical Engineering Plant Cost Index)
Distribution of TOC over CAPEX period	5 years: 10%, 30%, 25%, 20%, 15%
Working Capital	0
% of TOC being Depreciated	100%
Escalation of Operating Revenues and Cost	3% from average DoL Producer Price Index for Finished Goods

Table 5-17. Financial structure for commercial fuel projects.

	% of Total	Current $ cost	Current weighted cost
Debt	50	8%	4%
Equity	50	20%	10%
Total			14%

The cash flow analysis was used to determine the methanol RSP assuming a 5-year capital expenditure, a 35-year analysis period, a 30-year debt repayment period, a 20-year depreciation period, and a 20% ROE for the project. The details for both the reference case and the Sour PSA case can be found in Appendix Tables A5-15 and 16 respectively. Table 5-18 summarizes the results for comparison of both cases.

Table 5-18. Methanol RSP and project NPV results from the cash flow analysis for both the reference and the Sour PSA cases.

		Reference case	Sour PSA case
Methanol RSP	**($/Gal)**	1.89	1.82
	Fuel ($/Gal)	0.29	0.30
	FOC ($/Gal)	0.18	0.17
	VOC ($/Gal)	0.07	0.08
	Capital ($/Gal)	1.30	1.21
	CO2 TS&M ($/Gal)	0.05	0.06
NPV at discount rate ($x1000)			
	8%	3,652,241	3,332,799
	14%	974,214	888,975
	20%	0	0

As shown for IGCC applications, the significant capital savings on the scope of acid gas removal allowed by the use of the Sour PSA technology translate to a noticeably lower cost of product. Air Products Sour PSA technology reduces the cost of methanol by ~3.7% compared to incumbent technology.

Sensitivity of the cost of methanol

As was done for the IGCC cases, cash flow analysis was used to perform sensitivity analysis around the capital cost of the Sour PSA technology, the selling price of CO_2 at the plant's gate, and the cost of CO_2 emissions. The methanol RSP sensitivity to the capital cost of the Sour PSA technology was established by varying the total cost of the new technologies for the AGR scope (Sour PSA and AGE) from 50% to 200% of the value used to establish the RSP in Table 5-18. The resulting evolution of the methanol RSP is presented in Figure 5-14.

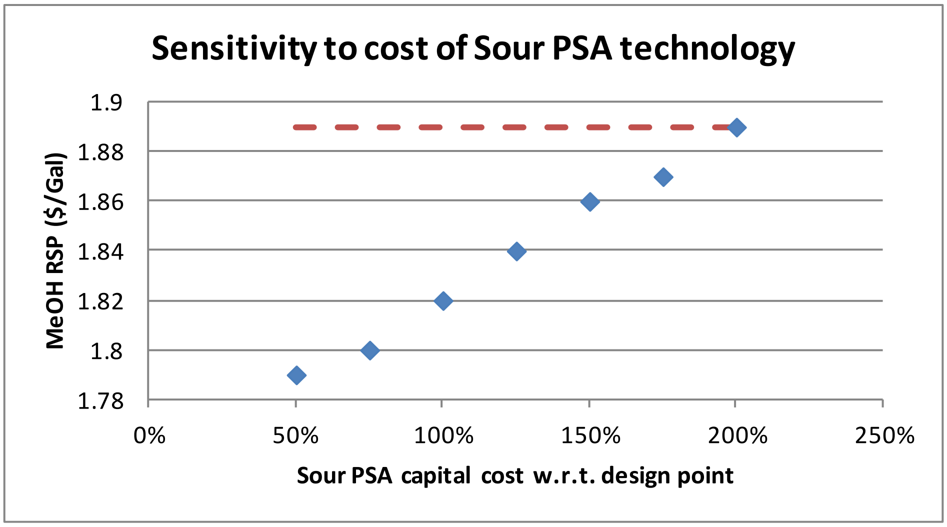


Figure 5-14. Evolution of the methanol RSP as a function of capital cost of the Sour PSA AGR technology.

The doted red line represents the methanol RSP for the reference case. This graph shows that the capital cost of the AGR based on Sour PSA technology (PSA and AGE) would have to be more than 2X higher than estimate used here to result in a higher RSP than the reference technology can achieve. This is a remarkable result given that this estimate already includes 20 % process and project contingencies for the Sour PSA.

The sensitivity of the COE to the performance of an AGR system based on the Sour PSA technology is more complex to establish. Indeed, lower performance of an AGR system means that products and/or recovery are not at the designed specifications. This can always be mitigated with larger-capacity or additional equipment and therefore would fall into the capital cost sensitivity presented above. Instead, it was decided to look at the sensitivity of the methanol RSP to the parasitic load of the AGR system since it is also a performance factor. Therefore, this study looked at the evolution of the COE when varying the power consumption of the entire AGR system (Sour PSA and AGE) from 50% to 200% of the design point used to establish the methanol RSP in Table 5-18. The resulting evolution of the methanol RSP is presented in Figure 5-15.

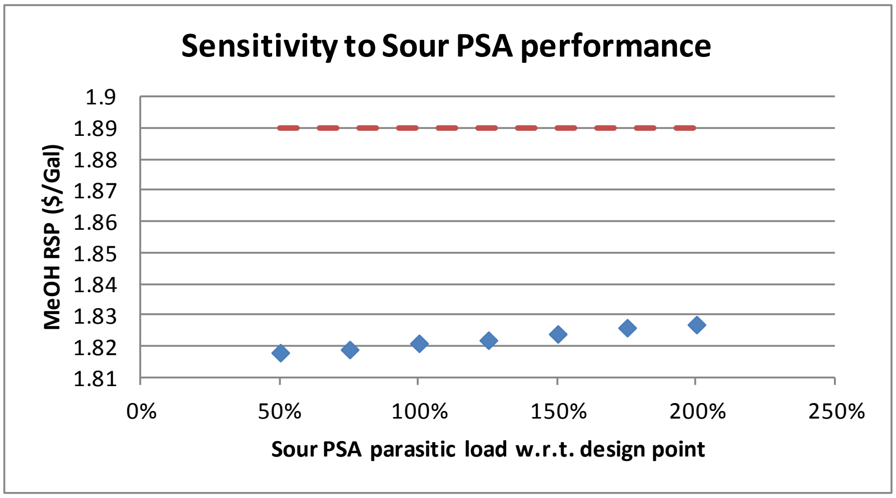


Figure 5-15. Evolution of the methanol RSP as a function of power consumption of the Sour PSA AGR technology.

The dotted red line represents the methanol RSP for the reference case. Again, this graph suggests that the power consumption would have to be more than 12 times higher for the Sour PSA to produce methanol at a higher cost than the reference case.

The evolution of the methanol RSP was then compared for both the reference and the Sour PSA cases as a function of the selling price of CO_2 at the gate. In that scenario, all captured CO_2 is sold at the gate, so no TS&M is incurred. The CO_2 price is established for the first year of operation (from 0 to 60 $/tonne) and increased by 3% every year according to the global economic assumptions in Table 5-16. Figure 5-16 suggests that, compared to the reference case, the methanol RSP in the Sour PSA case decreases faster as a function of increasing sale price of CO_2. This is the result of achieving higher CO_2 capture with the Sour PSA than the incumbent technology.

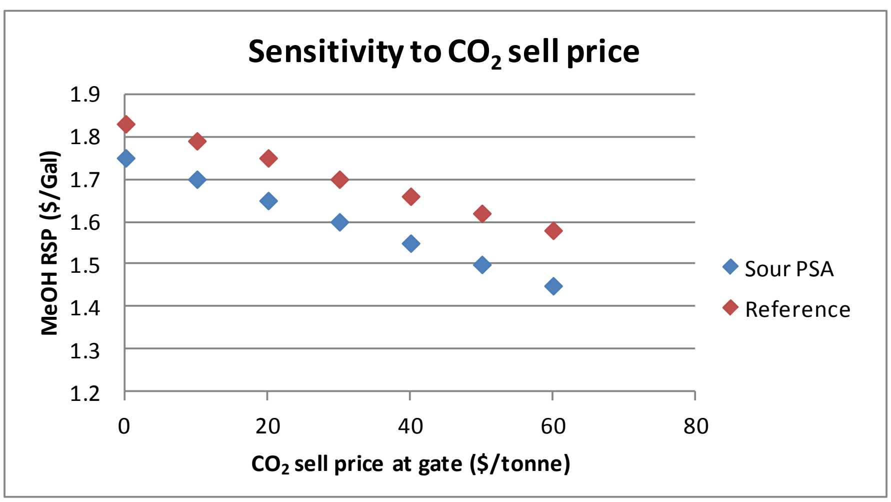


Figure 5-16: Evolution of the methanol RSP as a function of selling price of CO_2 at the gate for both Sour PSA and the reference technology.

Finally, Air Products looked at the sensitivity of the methanol RSP to a tax on the emitted CO_2. In that scenario, the tax is paid only on the CO_2 that is not captured by the AGR technology. There are two points of CO_2 emission in the plant: the exhaust from the gas turbine and the exhaust from the fired boiler in the methanol plant. The gas turbine CO_2 emissions are identical in both cases, and since they are independent of the AGR technology they were not accounted for in this sensitivity analysis (power could be imported). Therefore, the basis for comparison is only the CO_2 emissions from the coal-to-methanol process, again comparing the reference and Sour PSA cases. The tax is fixed at the first year of operation (from 0 to \$60/tone) and increased by 3% every year according to the global economic assumptions in Table 5-16. The resulting evolution of the COE is presented in Figure 5-17.

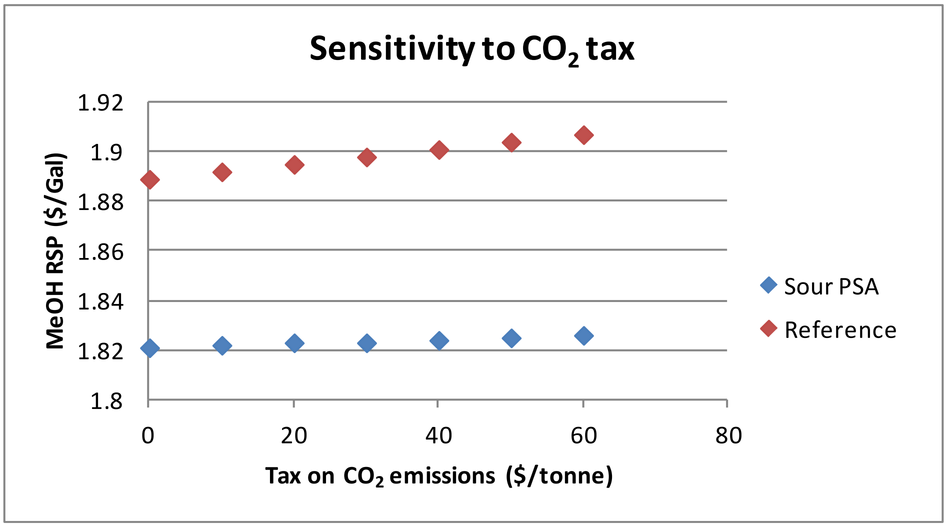


Figure 5-17. Evolution of the methanol RSP as a function of tax level on CO_2 emissions from the coal-to-methanol process only for both the reference and the Sour PSA cases.

Again, the results show that no matter the tax on emitted CO_2, the Sour PSA-based AGR technology always results in a lower methanol RSP than the incumbent technology. The methanol RSP is less sensitive to the CO_2 tax level in the case of Sour PSA technology mainly because of the higher level of CO_2 capture it can achieve.

Conclusion

Air Products has developed an acid gas removal technology based on adsorption (Sour PSA) that favorably compares with incumbent AGR technologies. During this DOE-sponsored study, Air Products has been able to increase the technology readiness level of the Sour PSA technology by successfully operating a two-bed test system on coal-derived sour syngas at the NCCC, concurrently validating the lifetime and performance of the adsorbent material. Both proprietary simulation and the data obtained during the testing at NCCC were used to further refine the estimate of the performance of the Sour PSA technology when scaled up to a commercial size. In-house experiments on sweet syngas combined with simulation work enabled Air Products to develop new PSA cycles that allow for further reduction in capital expenditure. Finally, the techno-economic analysis of the use the Sour PSA technology for both IGCC and coal-to-methanol application suggests significant benefit compared to incumbent AGR technologies. Indeed, this study showed a 4.6 % reduction in the cost of electricity and a 3.7% reduction in the cost of methanol when using Air Products' Sour PSA technology.

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Appendix

Table A5-1. Heat and mass balance for the reference case IGCC plant.

stream number	1	2	3	4	5	6	7	8	9	10
Mole Fraction										
H_2	0.14560						0.53255	0.82480	0.00470	
N_2	0.05240						0.06631	0.10340	0.00030	
CO	0.28300						0.01010	0.01560	0.00030	
CO_2	0.02570						0.38094	0.04600	0.99450	
CH_4	0.00000						0.00000	0.00010	0.00000	
Ar	0.00510						0.00640	0.01000	0.00020	
H_2S	0.00150						0.00210	0.00000	0.00000	
COS	0.00010						0.00000	0.00000	0.00000	
H_2O	0.48540						0.00160	0.00010	0.00000	
SO_2	0.00000						0.00000	0.00000	0.00000	
SO_3	0.00000						0.00000	0.00000	0.00000	
O_2	0.00000						0.00000	0.00000	0.00000	
NO	0.00000						0.00000	0.00000	0.00000	
NO_2	0.00000						0.00000	0.00000	0.00000	
Mole Flow Rate (lbmol/hr)	77885.00						61559.68	39126.00	21749.00	
Mass Flow Rate (lb./hr)	1546091.00	1546091.00	1546091.00	1546091.00	1546091.00	1546091.00	1251807.54	290459.69	952667.00	4206.00
Temperature (F)	450.00	875.30	400.00	527.40	315.71	95.00	95.00	420.00	162.00	347.00
Pressure (PSIA)	579.70	569.70	564.70	552.60	552.60	552.60	522.14	464.60	2214.70	
Enthalpy (Btu/lb)	-3317.69						-3203.63	-823.62	-3895.40	

Table A5-1. Heat and mass balance for the reference case IGCC plant (cont.).

stream number	11	12	13	14	15	16	17	18	19	20
Mole Fraction										
H_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N_2	0.01780	0.99210	0.75530	0.75530	0.75530	0.00000	0.00000	0.00000	0.00000	0.00000
CO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CO_2	0.00000	0.00000	0.00890	0.00890	0.00890	0.00000	0.00000	0.00000	0.00000	0.00000
CH_4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Ar	0.03180	0.00230	0.00900	0.00900	0.00900	0.00000	0.00000	0.00000	0.00000	0.00000
H_2S	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H_2O	0.00000	0.00020	0.12040	0.12040	0.12040	1.00000	1.00000	1.00000	1.00000	1.00000
SO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO_3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O_2	0.95040	0.00540	0.10640	0.10640	0.10640	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Mole Flow Rate (lbmol/hr)	102.55	36104.00	280029.46	280029.46	280029.46	40452.84	40452.84	39571.24	39526.94	39025.42
Mass Flow Rate (lb./hr)	3300.00	1013113.14	7696268.00	7696268.00	7696268.00	728758.00	728758.00	712875.89	712077.89	703042.90
Temperature (F)	90.00	385.00	1041.90	321.87	270.00	278.80	293.84	622.26	991.90	617.73
Pressure (PSIA)	125.00	384.00	15.00	15.00	15.00	2250.70	2250.70	1814.70	1814.70	476.70
Enthalpy (Btu/lb)	1.90	75.05	-250.42	-447.15	-460.72	-6618.55	-6603.29	-6105.66	-5395.34	-5559.81

Table A5-1. Heat and mass balance for the reference case IGCC plant (cont.).

stream number	21	22	23	24	25	26	27	28	29	30
Mole Fraction										
H_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH_4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Ar	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H_2S	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H_2O	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000
SO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO_3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Mole Flow Rate (lbmol/hr)	39025.42	2205.83	36745.32	39494.20	39494.20	39494.20	76239.52	620.09	3776.13	80635.74
Mass Flow Rate (lb./hr)	703042.90	39738.00	661966.90	711488.00	711488.00	711488.00	1373454.90	11170.99	68026.99	1452652.87
Temperature (F)	1000.00	843.13	483.47	278.80	298.02	409.00	90.06	572.24	48.00	92.20
Pressure (PSIA)	476.70	280.00	65.00	65.00	65.00	65.00	0.70	65.00	14.70	120.00
Enthalpy (Btu/lb)	-5349.08	-5425.93	-5596.13	-6622.85	-5690.68	-5633.15	-5913.98	-5552.41	-6854.82	-6810.37

Table A5-1. Heat and mass balance for the reference case IGCC plant (cont.).

stream number	31	32	33	34
Mole Fraction				
H_2	0.00000	0.00000	0.00000	0.00000
N_2	0.00000	0.77590	0.00000	0.00000
CO	0.00000	0.00000	0.00000	0.00000
CO_2	0.00000	0.00030	0.00000	0.00000
CH_4	0.00000	0.00000	0.00000	0.00000
Ar	0.00000	0.00930	0.00000	0.00000
H_2S	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000
H_2O	1.00000	0.00640	1.00000	1.00000
SO_2	0.00000	0.00000	0.00000	0.00000
SO_3	0.00000	0.00000	0.00000	0.00000
O_2	0.00000	0.20810	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000
NO_2	0.00000	0.00000	0.00000	0.00000
Mole Flow Rate (lbmol/hr)	80635.74	221239.00	881.61	480.66
Mass Flow Rate (lb./hr)	1452652.87	6392670.26	15882.11	8659.12
Temperature (F)	235.00	42.00	622.26	298.02
Pressure (PSIA)	110.00	13.00	1814.70	65.00
Enthalpy (Btu/lb)	-6667.31	-33.35	-6105.66	-5690.68

Table A5-2. Heat and mass balance for the Sour PSA case IGCC plant.

stream number	1	2	3	4	5	6	7	8	9	10
Mole Fraction										
H_2	0.14560						0.53255	0.84158	0.00791	
N_2	0.05240						0.06631	0.10271	0.00408	
CO	0.28300						0.01010	0.01516	0.00094	
CO_2	0.02570						0.38094	0.03063	0.98637	
CH_4	0.00000						0.00000	0.00000	0.00000	
Ar	0.00510						0.00640	0.00991	0.00045	
H_2S	0.00150						0.00210	0.00000	0.00025	
COS	0.00010						0.00000	0.00000	0.00000	
H_2O	0.48540						0.00160	0.00000	0.00000	
SO_2	0.00000						0.00000	0.00000	0.00000	
SO_3	0.00000						0.00000	0.00000	0.00000	
O_2	0.00000						0.00000	0.00000	0.00000	
NO	0.00000						0.00000	0.00000	0.00000	
NO_2	0.00000						0.00000	0.00000	0.00000	
Mole Flow Rate (lbmol/hr)	77885.00						61559.68	36106.29	22003.64	
Mass Flow Rate (lb./hr)	1546091.00	1546091.00	1546091.00	1546091.00	1546091.00	1546091.00	1251807.54	243452.68	959190.50	3858.72
Temperature (F)	450.00	875.30	400.00	527.40	292.00	95.00	95.00	95.00	162.00	347.00
Pressure (PSIA)	579.70	569.70	564.70	552.60	552.60	552.60	522.14	513.00	2215.00	
Enthalpy (Btu/lb)	-3317.69						-3203.63	-858.29	-3881.44	

Table A5-2. Heat and mass balance for the Sour PSA case IGCC plant (cont.).

stream number	11	12	13	14	15	16	17	18	19	20
Mole Fraction (%)										
H_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N_2	0.01780	0.99210	0.75530	0.75530	0.75530	0.00000	0.00000	0.00000	0.00000	0.00000
CO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CO_2	0.00000	0.00000	0.00890	0.00890	0.00890	0.00000	0.00000	0.00000	0.00000	0.00000
CH_4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Ar	0.03180	0.00230	0.00900	0.00900	0.00900	0.00000	0.00000	0.00000	0.00000	0.00000
H_2S	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H_2O	0.00000	0.00020	0.12040	0.12040	0.12040	1.00000	1.00000	1.00000	1.00000	1.00000
SO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO_3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O_2	0.95040	0.00540	0.10640	0.10640	0.10640	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Mole Flow Rate (lbmol/hr)	94.08	36597.30	282678.45	282678.45	282678.45	40998.17	40998.17	40104.69	40059.80	39551.51
Mass Flow Rate (lb./hr)	3027.52	1026955.50	7769072.30	7769072.30	7769072.30	738582.11	738582.11	722486.01	721677.25	712520.47
Temperature (F)	90.00	385.00	1041.90	321.87	270.00	278.80	294.79	622.26	991.90	617.73
Pressure (PSIA)	125.00	384.00	15.00	15.00	15.00	2250.70	2250.70	1814.70	1814.70	476.70
Enthalpy (Btu/lb)	1.90	75.05	-250.42	-447.15	-460.72	-6618.55	-6602.32	-6111.30	-5395.34	-5559.81

Table A5-2. Heat and mass balance for the Sour PSA case IGCC plant (cont.).

stream number	21	22	23	24	25	26	27	28	29	30
Mole Fraction (%)										
H_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH_4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Ar	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H_2S	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H_2O	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000
SO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO_3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Mole Flow Rate (lbmol/hr)	39551.51	2205.83	37270.35	40026.61	40026.61	40026.61	77296.96	628.51	3828.50	81753.96
Mass Flow Rate (lb./hr)	712520.47	39738.00	671425.35	721079.30	721079.30	721079.30	1392504.65	11322.66	68970.36	1472797.66
Temperature (F)	1000.00	843.13	483.47	278.80	298.02	409.00	90.06	572.24	48.00	92.20
Pressure (PSIA)	476.70	280.00	65.00	65.00	65.00	65.00	0.70	65.00	14.70	120.00

Table A5-2. Heat and mass balance for the Sour PSA case IGCC plant (cont.).

stream number	31	32	33	34	35	36	37	38	39	40
Mole Fraction (%)										
H_2	0.00000	0.00000	0.00000	0.00000	0.09419	0.09502	0.00000	0.00000	0.00000	0.00000
N_2	0.00000	0.77590	0.00000	0.00000	0.01467	0.01480	0.00000	0.00000	0.00000	0.00000
CO	0.00000	0.00000	0.00000	0.00000	0.00292	0.00295	0.00000	0.00000	0.00000	0.00000
CO_2	0.00000	0.00030	0.00000	0.00000	0.87786	0.88559	0.70236	0.00000	0.99601	0.00000
CH_4	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Ar	0.00000	0.00930	0.00000	0.00000	0.00142	0.00143	0.00000	0.00000	0.00000	0.00000
H_2S	0.00000	0.00000	0.00000	0.00000	0.00508	0.00022	0.16413	0.00000	0.00002	0.00000
COS	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H_2O	1.00000	0.00640	1.00000	1.00000	0.00387	0.00000	0.13351	1.00000	0.00397	1.00000
SO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO_3	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O_2	0.00000	0.20810	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO_2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Mole Flow Rate (lbmol/hr)	81753.96	223413.08	893.48	481.61	25453.39	25231.19	753.48		531.28	195.21
Mass Flow Rate (lb./hr)	1472797.66	6455490.12	16096.10	8676.14	1008354.85	1002364.50	29316.41		23326.07	3516.62
Temperature (F)	235.00	42.00	622.26	298.02	95.00	87.65	87.65		105.16	111.20
Pressure (PSIA)	110.00	13.00	1814.70	65.00	17.64	17.64	17.64		17.64	18.85
Enthalpy (Btu/lb)	-6667.31	-33.35	-6111.30	-5690.68	-3762.74	-3775.57	-3496.75		-3844.63	-6792.17

Table A5-2. Heat and mass balance for the Sour PSA case IGCC plant (cont.).

stream number	41	42	43	44
Mole Fraction (%)				
H_2	0.00000	0.00791	0.68887	0.82905
N_2	0.00000	0.00408	0.08785	0.10149
CO	0.00000	0.00094	0.01662	0.01528
CO_2	0.00000	0.98637	0.19847	0.04440
CH_4	0.00000	0.00000	0.00000	0.00000
Ar	0.00000	0.00045	0.00817	0.00977
H_2S	0.00000	0.00025	0.00002	0.00000
COS	0.00000	0.00000	0.00000	0.00000
H_2O	1.00000	0.00000	0.00000	0.00000
SO_2	0.00000	0.00000	0.00000	0.00000
SO_3	0.00000	0.00000	0.00000	0.00000
O_2	0.00000	0.00000	0.00000	0.00000
NO	0.00000	0.00000	0.00000	0.00000
NO_2	0.00000	0.00000	0.00000	0.00000
Mole Flow Rate (lbmol/hr)	193.10	22003.64	3227.55	39333.84
Mass Flow Rate (lb./hr)	3478.63	959190.50	43174.00	286626.68
Temperature (F)	111.20	59.19	56.99	420.00
Pressure (PSIA)	18.85	240.52	517.06	464.60
Enthalpy (Btu/lb)	-6791.16	-3842.97	-2586.48	-798.22

Table A5-3. Total plant cost summary for the reference IGCC case.

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Eng. CM H.O. & Fee	Process Contingencies %	Process Contingencies $ x1000	Project Contingencies %	Project Contingencies $ x1000	TOTAL PLANT COST $ x1000	TOTAL PLANT COST $/kW
1	**COAL & SORBENT HANDLING**	**$17,736**	**$3,296**	**$13,755**	**$0**	**$0**	**$34,788**	**$3,157**	**0%**	**$0**	**20%**	**$7,589**	**$45,533**	**$102**
1.1	Coal Receive & Unload	$4,658	$0	$2,276	$0	$0	$6,934	$621	0%	$0	20%	$1,511	$9,066	$20
1.2	Coal Stackout & Reclaim	$6,019	$0	$1,459	$0	$0	$7,478	$655	0%	$0	20%	$1,627	$9,760	$22
1.3	Coal Conveyors & Yd Crush	$5,596	$0	$1,444	$0	$0	$7,040	$618	0%	$0	20%	$1,531	$9,189	$21
1.4	Other Coal Handling	$1,464	$0	$334	$0	$0	$1,798	$157	0%	$0	20%	$391	$2,346	$5
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.9	Coal & Sorbent Hnd.Foundations	$0	$3,296	$8,241	$0	$0	$11,538	$1,106	0%	$0	20%	$2,529	$15,172	$34
2	**COAL & SORBENT PREP & FEED**	**$133,727**	**$11,142**	**$22,817**	**$0**	**$0**	**$167,686**	**$14,548**	**0%**	**$0**	**20%**	**$36,447**	**$218,680**	**$491**
2.1	Coal Crushing & Drying	$53,837	$3,234	$7,845	$0	$0	$64,915	$5,602	0%	$0	20%	$14,103	$84,620	$190
2.2	Prepared Coal Storage & Feed	$2,314	$554	$363	$0	$0	$3,232	$276	0%	$0	20%	$702	$4,210	$9
2.3	Dry Coal Injection System	$76,173	$884	$7,074	$0	$0	$84,132	$7,246	0%	$0	20%	$18,276	$109,653	$246
2.4	Misc.Coal Prep & Feed	$1,402	$1,020	$3,060	$0	$0	$5,482	$504	0%	$0	20%	$1,197	$7,184	$16
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$5,450	$4,475	$0	$0	$9,925	$920	0%	$0	20%	$2,169	$13,013	$29
3	**FEEDWATER & MISC. BOP SYSTEMS**	**$8,241**	**$6,034**	**$8,474**	**$0**	**$0**	**$22,750**	**$2,153**	**0%**	**$0**	**23%**	**$5,847**	**$30,750**	**$69**
3.1	Feedwater System	$1,974	$3,390	$1,789	$0	$0	$7,153	$662	0%	$0	20%	$1,563	$9,379	$21
3.2	Water Makeup & Pretreating	$616	$65	$344	$0	$0	$1,026	$98	0%	$0	30%	$337	$1,461	$3
3.3	Other Feedwater Subsystems	$1,080	$365	$329	$0	$0	$1,774	$159	0%	$0	20%	$387	$2,320	$5
3.4	Service Water Systems	$352	$726	$2,519	$0	$0	$3,597	$351	0%	$0	30%	$1,185	$5,133	$12
3.5	Other Boiler Plant Systems	$1,892	$732	$1,816	$0	$0	$4,440	$421	0%	$0	20%	$972	$5,834	$13
3.6	FO Supply Sys & Nat Gas	$319	$603	$562	$0	$0	$1,484	$143	0%	$0	20%	$325	$1,952	$4
3.7	Waste Treatment Equipment	$861	$0	$525	$0	$0	$1,386	$135	0%	$0	30%	$456	$1,977	$4
3.8	Misc. Power Plant Equipment	$1,147	$154	$590	$0	$0	$1,891	$183	0%	$0	30%	$622	$2,696	$6
4	**GASIFIER & ACCESSORIES**	**$361,748**	**$14,291**	**$78,221**	**$0**	**$0**	**$454,261**	**$42,467**	**6%**	**$27,564**	**14%**	**$72,023**	**$596,315**	**$1,339**
4.1	Gasifier, Syngas Cooler & Auxiliaries (Siemens)	$125,254	$0	$58,508	$0	$0	$183,762	$16,327	15%	$27,564	15%	$34,148	$261,801	$588
4.2	Syngas Cooling	w/4.1	$0	w/4.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.3	ASU/Oxidant Compression	$205,287	$0	w/equip	$0	$0	$205,287	$19,899	0%	$0	10%	$22,519	$247,704	$556
4.4	LT Heat Recovery & FG Saturation	$31,207	$0	$11,863	$0	$0	$43,070	$4,204	0%	$0	20%	$9,455	$56,729	$127
4.5	Misc. Gasification Equipment	w/4.1 &4.2	$0	w/4.1&4.2	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.6	Flare Stack System	$0	$1,862	$758	$0	$0	$2,620	$251	0%	$0	20%	$574	$3,445	$8
4.8	Major Component Rigging	w/4.1 &4.2	$0	w/4.1&4.2	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.9	Gasification Foundations	$0	$12,430	$7,092	$0	$0	$19,522	$1,787	0%	$0	25%	$5,327	$26,636	$60
5A	**GAS CLEANUP & PIPING**	**$93,900**	**$3,199**	**$79,858**	**$0**	**$0**	**$176,957**	**$17,094**	**17%**	**$29,230**	**20%**	**$44,844**	**$268,125**	**$602**
5A.1	Double Stage Selexol	$78,695	$0	$66,775	$0	$0	$145,470	$14,068	20%	$29,094	20%	$37,726	$226,358	$508
5A.2	Elemental Sulfur Plant	$5,649	$1,126	$7,287	$0	$0	$14,062	$1,366	0%	$0	20%	$3,085	$18,513	$42
5A.3	Mercury Removal	$1,543	$0	$1,175	$0	$0	$2,718	$262	5%	$136	20%	$623	$3,739	$8
5A.4	Shift Reactors	$8,014	$0	$3,226	$0	$0	$11,240	$1,078	0%	$0	20%	$2,464	$14,782	$33
5A.5	Particulate Removal	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
5A.6	Blowback Gas Systems	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
5A.7	Fuel Gas Piping	$0	$1,030	$721	$0	$0	$1,751	$163	0%	$0	20%	$383	$2,297	$5
5A.9	HGCU Foundations	$0	$1,043	$673	$0	$0	$1,717	$157	0%	$0	30%	$562	$2,436	$5
5B	**CO2 COMPRESSION**	**$20,740**	**$0**	**$11,773**	**$0**	**$0**	**$32,513**	**$3,129**	**0%**	**$0**	**20%**	**$7,129**	**$42,771**	**$96**
5B.1	CO2 Removal System	w/5A.1	$0	w/5A.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
5B.2	CO2 Compression & Drying	$20,740	$0	$11,773	$0	$0	$32,513	$3,129	0%	$0	20%	$7,129	$42,771	$96
6	**COMBUSTION TURBINE/ACCESSORIES**	**$102,589**	**$899**	**$8,333**	**$0**	**$0**	**$111,820**	**$10,598**	**10%**	**$10,993**	**10%**	**$13,755**	**$147,166**	**$330**
6.1	Combustion Turbine Generator	$102,589	$0	$7,339	$0	$0	$109,927	$10,421	10%	$10,993	10%	$13,134	$144,475	$324
6.2	Open	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$899	$994	$0	$0	$1,893	$177	0%	$0	30%	$621	$2,691	$6
7	**HRSG, DUCTING & STACK**	**$38,933**	**$2,801**	**$8,683**	**$0**	**$0**	**$50,418**	**$4,769**	**0%**	**$0**	**11%**	**$6,235**	**$61,422**	**$138**
7.1	Heat Recovery Steam Generator	$35,005	$0	$4,977	$0	$0	$39,982	$3,801	0%	$0	10%	$4,378	$48,162	$108
7.2	Open	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
7.3	Ductwork	$0	$2,014	$1,474	$0	$0	$3,488	$307	0%	$0	20%	$759	$4,554	$10
7.4	Stack	$3,928	$0	$1,476	$0	$0	$5,404	$517	0%	$0	10%	$592	$6,514	$15
7.9	HRSG,Duct & Stack Foundations	$0	$787	$756	$0	$0	$1,543	$144	0%	$0	30%	$506	$2,193	$5
8	**STEAM TURBINE GENERATOR**	**$61,663**	**$894**	**$15,518**	**$0**	**$0**	**$78,075**	**$7,568**	**0%**	**$0**	**16%**	**$13,855**	**$99,499**	**$223**
8.1	Steam TG & Accessories	$26,169	$0	$4,364	$0	$0	$30,534	$2,930	0%	$0	10%	$3,346	$36,809	$83
8.2	Turbine Plant Auxiliaries	$181	$0	$414	$0	$0	$594	$58	0%	$0	10%	$65	$717	$2
8.3a	Condenser & Auxiliaries	$3,127	$0	$999	$0	$0	$4,126	$395	0%	$0	10%	$452	$4,972	$11
8.3b	Air Cooled Condenser	$28,654	$0	$5,744	$0	$0	$34,398	$3,440	0%	$0	20%	$7,568	$45,406	$102
8.4	Steam Piping	$3,533	$0	$2,485	$0	$0	$6,018	$517	0%	$0	25%	$1,634	$8,168	$18
8.9	TG Foundations	$0	$894	$1,512	$0	$0	$2,406	$229	0%	$0	30%	$790	$3,424	$8
9	**COOLING WATER SYSTEM**	**$8,744**	**$8,581**	**$7,303**	**$0**	**$0**	**$24,629**	**$2,288**	**0%**	**$0**	**20%**	**$5,511**	**$32,427**	**$73**
9.1	Cooling Towers	$6,023	$0	$1,096	$0	$0	$7,119	$678	0%	$0	15%	$1,170	$8,966	$20
9.2	Circulating Water Pumps	$1,566	$0	$100	$0	$0	$1,667	$140	0%	$0	15%	$271	$2,078	$5
9.3	Circ.Water System Auxiliaries	$136	$0	$19	$0	$0	$155	$14	0%	$0	15%	$25	$195	$0
9.4	Circ.Water Piping	$0	$5,693	$1,476	$0	$0	$7,169	$648	0%	$0	20%	$1,563	$9,380	$21
9.5	Make-up Water System	$347	$0	$496	$0	$0	$843	$81	0%	$0	20%	$185	$1,109	$2
9.6	Component Cooling Water Sys	$672	$804	$572	$0	$0	$2,048	$192	0%	$0	20%	$448	$2,687	$6
9.9	Circ.Water System Foundations	$0	$2,085	$3,544	$0	$0	$5,628	$534	0%	$0	30%	$1,849	$8,011	$18
10	**ASH/SPENT SORBENT HANDLING SYS**	**$23,855**	**$1,642**	**$11,829**	**$0**	**$0**	**$37,326**	**$3,582**	**0%**	**$0**	**11%**	**$4,430**	**$45,338**	**$102**
10.1	Slag Dewatering & Cooling	$21,097	$0	$10,404	$0	$0	$31,501	$3,027	0%	$0	10%	$3,453	$37,981	$85
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.4	High Temperature Ash Piping	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.5	Other Ash Recovery Equipment	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.6	Ash Storage Silos	$625	$0	$680	$0	$0	$1,305	$127	0%	$0	15%	$215	$1,647	$4
10.7	Ash Transport & Feed Equipment	$838	$0	$202	$0	$0	$1,040	$97	0%	$0	15%	$171	$1,308	$3
10.8	Misc. Ash Handling Equipment	$1,294	$1,586	$474	$0	$0	$3,354	$319	0%	$0	15%	$551	$4,224	$9
10.9	Ash/Spent Sorbent Foundation	$0	$56	$69	$0	$0	$125	$12	0%	$0	30%	$41	$178	$0
11	**ACCESSORY ELECTRIC PLANT**	**$34,245**	**$13,752**	**$26,645**	**$0**	**$0**	**$74,642**	**$6,420**	**0%**	**$0**	**19%**	**$15,438**	**$96,500**	**$217**
11.1	Generator Equipment	$970	$0	$960	$0	$0	$1,930	$184	0%	$0	10%	$211	$2,325	$5
11.2	Station Service Equipment	$5,166	$0	$466	$0	$0	$5,632	$519	0%	$0	10%	$615	$6,766	$15
11.3	Switchgear & Motor Control	$9,551	$0	$1,737	$0	$0	$11,288	$1,047	0%	$0	15%	$1,850	$14,185	$32
11.4	Conduit & Cable Tray	$0	$4,437	$14,637	$0	$0	$19,074	$1,845	0%	$0	25%	$5,230	$26,148	$59
11.5	Wire & Cable	$0	$8,477	$5,569	$0	$0	$14,046	$1,020	0%	$0	25%	$3,767	$18,833	$42
11.6	Protective Equipment	$0	$681	$2,480	$0	$0	$3,161	$309	0%	$0	15%	$521	$3,991	$9
11.7	Standby Equipment	$243	$0	$237	$0	$0	$480	$46	0%	$0	15%	$79	$605	$1
11.8	Main Power Transformers	$18,315	$0	$145	$0	$0	$18,459	$1,396	0%	$0	15%	$2,978	$22,833	$51
11.9	Electrical Foundations	$0	$157	$414	$0	$0	$571	$55	0%	$0	30%	$188	$813	$2
12	**INSTRUMENTATION & CONTROL**	**$11,984**	**$2,205**	**$7,721**	**$0**	**$0**	**$21,910**	**$1,987**	**5%**	**$1,095**	**17%**	**$4,164**	**$29,156**	**$65**
12.1	IGCC Control Equipment	w/4.1	$0	w/4.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.2	Combustion Turbine Control	w/6.1	$0	w/6.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.4	Other Major Component Control	$1,183	$0	$790	$0	$0	$1,973	$187	5%	$99	15%	$339	$2,598	$6
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.6	Control Boards,Panels & Racks	$272	$0	$174	$0	$0	$446	$42	5%	$22	20%	$102	$613	$1
12.7	Computer & Accessories	$6,311	$0	$202	$0	$0	$6,512	$598	5%	$326	10%	$744	$8,179	$18
12.8	Instrument Wiring & Tubing	$0	$2,205	$4,507	$0	$0	$6,712	$570	5%	$336	25%	$1,904	$9,522	$21
12.9	Other I & C Equipment	$4,218	$0	$2,048	$0	$0	$6,266	$590	5%	$313	15%	$1,075	$8,245	$19
13	**IMPROVEMENTS TO SITE**	**$3,626**	**$2,137**	**$8,946**	**$0**	**$0**	**$14,709**	**$1,453**	**0%**	**$0**	**30%**	**$4,849**	**$21,011**	**$47**
13.1	Site Preparation	$0	$114	$2,431	$0	$0	$2,545	$253	0%	$0	30%	$839	$3,637	$8
13.2	Site Improvements	$0	$2,023	$2,689	$0	$0	$4,712	$465	0%	$0	30%	$1,553	$6,730	$15
13.3	Site Facilities	$3,626	$0	$3,826	$0	$0	$7,452	$735	0%	$0	30%	$2,456	$10,643	$24
14	**BUILDINGS & STRUCTURES**	**$0**	**$6,999**	**$7,884**	**$0**	**$0**	**$14,882**	**$1,353**	**0%**	**$0**	**16%**	**$2,670**	**$18,906**	**$42**
14.1	Combustion Turbine Area	$0	$295	$167	$0	$0	$463	$40	0%	$0	20%	$101	$603	$1
14.2	Steam Turbine Building	$0	$2,272	$3,236	$0	$0	$5,508	$507	0%	$0	15%	$902	$6,918	$16
14.3	Administration Building	$0	$958	$695	$0	$0	$1,652	$147	0%	$0	15%	$270	$2,069	$5
14.4	Circulation Water Pumphouse	$0	$176	$94	$0	$0	$270	$23	0%	$0	15%	$44	$337	$1
14.5	Water Treatment Buildings	$0	$515	$503	$0	$0	$1,018	$91	0%	$0	15%	$166	$1,276	$3
14.6	Machine Shop	$0	$490	$336	$0	$0	$826	$74	0%	$0	15%	$135	$1,035	$2
14.7	Warehouse	$0	$791	$511	$0	$0	$1,302	$115	0%	$0	15%	$213	$1,629	$4
14.8	Other Buildings & Structures	$0	$464	$361	$0	$0	$825	$74	0%	$0	20%	$180	$1,078	$2
14.9	Waste Treating Building & Str.	$0	$1,037	$1,982	$0	$0	$3,019	$282	0%	$0	20%	$660	$3,961	$9
	TOTAL COST	**$921,733**	**$77,873**	**$317,760**	**$0**	**$0**	**$1,317,366**	**$122,564**	**5%**	**$68,882**	**16%**	**$244,786**	**$1,753,599**	**$3,938**

Table A5-4. Total plant cost summary for the Sour PSA IGCC case.

Acct.		Equipment	Material	Labor		Sales	Bare Erected	Eng. CM	Process Contingencies		Project Contingencies		TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost	H.O. & Fee	%	$ x1000	%	$ x1000	$ x1000	$/kW
1	**COAL & SORBENT HANDLING**	$17,736	$3,296	$13,755	$0	$0	$34,788	$3,157	0%	$0	20%	$7,589	$45,533	$101
1.1	Coal Receive & Unload	$4,658	$0	$2,276	$0	$0	$6,934	$621	0%	$0	20%	$1,511	$9,066	$20
1.2	Coal Stackout & Reclaim	$6,019	$0	$1,459	$0	$0	$7,478	$655	0%	$0	20%	$1,627	$9,760	$22
1.3	Coal Conveyors & Yd Crush	$5,596	$0	$1,444	$0	$0	$7,040	$618	0%	$0	20%	$1,531	$9,189	$20
1.4	Other Coal Handling	$1,464	$0	$334	$0	$0	$1,798	$157	0%	$0	20%	$391	$2,346	$5
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.9	Coal & Sorbent Hnd.Foundations	$0	$3,296	$8,241	$0	$0	$11,538	$1,106	0%	$0	20%	$2,529	$15,172	$34
2	**COAL & SORBENT PREP & FEED**	$133,727	$11,142	$22,817	$0	$0	$167,686	$14,548	0%	$0	20%	$36,447	$218,680	$487
2.1	Coal Crushing & Drying	$53,837	$3,234	$7,845	$0	$0	$64,915	$5,602	0%	$0	20%	$14,103	$84,620	$189
2.2	Prepared Coal Storage & Feed	$2,314	$554	$363	$0	$0	$3,232	$276	0%	$0	20%	$702	$4,210	$9
2.3	Dry Coal Injection System	$76,173	$884	$7,074	$0	$0	$84,132	$7,246	0%	$0	20%	$18,276	$109,653	$244
2.4	Misc.Coal Prep & Feed	$1,402	$1,020	$3,060	$0	$0	$5,482	$504	0%	$0	20%	$1,197	$7,184	$16
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$5,450	$4,475	$0	$0	$9,925	$920	0%	$0	20%	$2,169	$13,013	$29
3	**FEEDWATER & MISC. BOP SYSTEMS**	$8,241	$6,034	$8,474	$0	$0	$22,750	$2,153	0%	$0	23%	$5,847	$30,750	$69
3.1	Feedwater System	$1,974	$3,390	$1,789	$0	$0	$7,153	$662	0%	$0	20%	$1,563	$9,379	$21
3.2	Water Makeup & Pretreating	$616	$65	$344	$0	$0	$1,026	$98	0%	$0	30%	$337	$1,461	$3
3.3	Other Feedwater Subsystems	$1,080	$365	$329	$0	$0	$1,774	$159	0%	$0	20%	$387	$2,320	$5
3.4	Service Water Systems	$352	$726	$2,519	$0	$0	$3,597	$351	0%	$0	30%	$1,185	$5,133	$11
3.5	Other Boiler Plant Systems	$1,892	$732	$1,816	$0	$0	$4,440	$421	0%	$0	20%	$972	$5,834	$13
3.6	FO Supply Sys & Nat Gas	$319	$603	$562	$0	$0	$1,484	$143	0%	$0	20%	$325	$1,952	$4
3.7	Waste Treatment Equipment	$861	$0	$525	$0	$0	$1,386	$135	0%	$0	30%	$456	$1,977	$4
3.8	Misc. Power Plant Equipment	$1,147	$154	$590	$0	$0	$1,891	$183	0%	$0	30%	$622	$2,696	$6
4	**GASIFIER & ACCESSORIES**	$363,524	$14,291	$78,256	$0	$0	$456,071	$42,642	6%	$27,564	14%	$72,227	$598,505	$1,334
4.1	Gasifier, Syngas Cooler & Auxiliaries (Siemens)	$125,254	$0	$58,508	$0	$0	$183,762	$16,327	15%	$27,564	15%	$34,148	$261,801	$584
4.2	Syngas Cooling	w/4.1	$0	w/4.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.3	ASU/Oxidant Compression	$207,063	$0	w/equip	$0	$0	$207,063	$20,071	0%	$0	10%	$22,713	$249,847	$557
4.4	LT Heat Recovery & FG Saturation	$31,207	$0	$11,863	$0	$0	$43,070	$4,204	0%	$0	20%	$9,455	$56,729	$126
4.5	Misc. Gasification Equipment	w/4.1 &4.2	$0	w/4.1&4.2	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.6	Flare Stack System	$0	$1,862	$758	$0	$0	$2,620	$251	0%	$0	20%	$574	$3,445	$8
4.8	Major Component Rigging	w/4.1 &4.2	$0	w/4.1&4.2	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.9	Gasification Foundations	$0	$12,430	$7,127	$0	$0	$19,557	$1,790	0%	$0	25%	$5,337	$26,684	$59
5A	**GAS CLEANUP & PIPING**	$72,077	$2,923	$46,262	$0	$0	$121,262	$11,740	14%	$16,701	19%	$28,225	$177,929	$397
5A.1	Sour PSA System	$23,942	$0	$20,316	$0	$0	$44,258	$4,297	20%	$8,852	20%	$11,481	$68,888	$154
5A.2	Elemental Sulfur Plant	$5,478	$1,092	$7,067	$0	$0	$13,637	$1,324	0%	$0	20%	$2,992	$17,954	$40
5A.3	Mercury Removal	$1,543	$0	$1,175	$0	$0	$2,718	$262	5%	$136	20%	$623	$3,739	$8
5A.4	Shift Reactors	$8,014	$0	$3,226	$0	$0	$11,240	$1,078	0%	$0	20%	$2,464	$14,782	$33
5A.5	Particulate Removal	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
5A.6	Blowback Gas Systems	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
5A.7	Fuel Gas Piping	$0	$1,030	$721	$0	$0	$1,751	$163	0%	$0	20%	$383	$2,297	$5
5A.8	AGE	$11,100	$0	$4,440	$0	$0	$15,540	$1,507	10%	$1,554	10%	$1,860	$20,462	$46
5A.8b	Inert Rejection Unit	$22,000	$0	$8,800	$0	$0	$30,800	$2,988	20%	$6,160	20%	$7,990	$47,937	$107
5A9	Sour Gas Oxycombustor	$0	$0	$0	$0	$0	$0	$0	20%	$0	20%	$0	$0	$0
5A.10	HGCU Foundations	$0	$801	$517	$0	$0	$1,318	$121	0%	$0	30%	$432	$1,870	$4
5B	**CO2 COMPRESSION**	$13,501	$0	$7,664	$0	$0	$21,164	$2,037	0%	$0	20%	$4,640	$27,841	$62
5B.1	CO2 Removal System	w/5A.8	$0	w/5A.8	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
5B.2	CO2 Compression & Drying	$13,501	$0	$7,664	$0	$0	$21,164	$2,037	0%	$0	20%	$4,640	$27,841	$62
6	**COMBUSTION TURBINE/ACCESSORIES**	$103,181	$904	$8,381	$0	$0	$112,465	$10,659	10%	$11,056	10%	$13,834	$148,015	$330
6.1	Combustion Turbine Generator	$103,181	$0	$7,381	$0	$0	$110,562	$10,481	10%	$11,056	10%	$13,210	$145,309	$324
6.2	Open	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$904	$1,000	$0	$0	$1,904	$178	0%	$0	30%	$625	$2,707	$6
7	**HRSG, DUCTING & STACK**	$39,116	$2,816	$8,725	$0	$0	$50,657	$4,792	0%	$0	11%	$6,265	$61,714	$138
7.1	Heat Recovery Steam Generator	$35,168	$0	$5,001	$0	$0	$40,169	$3,819	0%	$0	10%	$4,399	$48,387	$108
7.2	Open	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
7.3	Ductwork	$0	$2,025	$1,481	$0	$0	$3,506	$308	0%	$0	20%	$763	$4,577	$10
7.4	Stack	$3,948	$0	$1,483	$0	$0	$5,432	$520	0%	$0	10%	$595	$6,547	$15
7.9	HRSG,Duct & Stack Foundations	$0	$791	$760	$0	$0	$1,551	$145	0%	$0	30%	$509	$2,204	$5
8	**STEAM TURBINE GENERATOR**	$62,069	$900	$15,620	$0	$0	$78,589	$7,618	0%	$0	16%	$13,947	$100,154	$223
8.1	Steam TG & Accessories	$26,340	$0	$4,393	$0	$0	$30,732	$2,949	0%	$0	10%	$3,368	$37,049	$83
8.2	Turbine Plant Auxiliaries	$182	$0	$416	$0	$0	$598	$58	0%	$0	10%	$66	$722	$2
8.3a	Condenser & Auxiliaries	$3,148	$0	$1,005	$0	$0	$4,153	$397	0%	$0	10%	$455	$5,005	$11
8.3b	Air Cooled Condenser	$28,843	$0	$5,782	$0	$0	$34,626	$3,463	0%	$0	20%	$7,618	$45,706	$102
8.4	Steam Piping	$3,556	$0	$2,501	$0	$0	$6,058	$521	0%	$0	25%	$1,645	$8,223	$18
8.9	TG Foundations	$0	$900	$1,522	$0	$0	$2,422	$230	0%	$0	30%	$796	$3,447	$8
9	**COOLING WATER SYSTEM**	$9,075	$8,723	$7,460	$0	$0	$25,258	$2,347	0%	$0	20%	$5,631	$33,236	$74
9.1	Cooling Towers	$6,337	$0	$1,153	$0	$0	$7,490	$713	0%	$0	15%	$1,231	$9,434	$21
9.2	Circulating Water Pumps	$1,576	$0	$101	$0	$0	$1,677	$141	0%	$0	15%	$273	$2,092	$5
9.3	Circ.Water System Auxiliaries	$137	$0	$19	$0	$0	$156	$15	0%	$0	15%	$26	$196	$0
9.4	Circ.Water Piping	$0	$5,726	$1,484	$0	$0	$7,210	$651	0%	$0	20%	$1,572	$9,434	$21
9.5	Make-up Water System	$349	$0	$499	$0	$0	$847	$82	0%	$0	20%	$186	$1,115	$2
9.6	Component Cooling Water Sys	$676	$901	$641	$0	$0	$2,218	$208	0%	$0	20%	$485	$2,911	$6
9.9	Circ.Water System Foundations	$0	$2,096	$3,563	$0	$0	$5,659	$537	0%	$0	30%	$1,859	$8,054	$18
10	**ASH/SPENT SORBENT HANDLING SYS**	$23,855	$1,642	$11,829	$0	$0	$37,326	$3,582	0%	$0	11%	$4,430	$45,338	$101
10.1	Slag Dewatering & Cooling	$21,097	$0	$10,404	$0	$0	$31,501	$3,027	0%	$0	10%	$3,453	$37,981	$85
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.4	High Temperature Ash Piping	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.5	Other Ash Recovery Equipment	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.6	Ash Storage Silos	$625	$0	$680	$0	$0	$1,305	$127	0%	$0	15%	$215	$1,647	$4
10.7	Ash Transport & Feed Equipment	$838	$0	$202	$0	$0	$1,040	$97	0%	$0	15%	$171	$1,308	$3
10.8	Misc. Ash Handling Equipment	$1,294	$1,586	$474	$0	$0	$3,354	$319	0%	$0	15%	$551	$4,224	$9
10.9	Ash/Spent Sorbent Foundation	$0	$56	$69	$0	$0	$125	$12	0%	$0	30%	$41	$178	$0
11	**ACCESSORY ELECTRIC PLANT**	$33,396	$13,836	$26,790	$0	$0	$74,022	$6,378	0%	$0	19%	$15,360	$95,760	$213
11.1	Generator Equipment	$975	$0	$965	$0	$0	$1,939	$185	0%	$0	10%	$212	$2,337	$5
11.2	Station Service Equipment	$5,199	$0	$469	$0	$0	$5,668	$523	0%	$0	10%	$619	$6,810	$15
11.3	Switchgear & Motor Control	$9,613	$0	$1,748	$0	$0	$11,361	$1,053	0%	$0	15%	$1,862	$14,276	$32
11.4	Conduit & Cable Tray	$0	$4,465	$14,731	$0	$0	$19,196	$1,857	0%	$0	25%	$5,263	$26,316	$59
11.5	Wire & Cable	$0	$8,531	$5,605	$0	$0	$14,136	$1,027	0%	$0	25%	$3,791	$18,953	$42
11.6	Protective Equipment	$0	$681	$2,480	$0	$0	$3,161	$309	0%	$0	15%	$521	$3,991	$9
11.7	Standby Equipment	$244	$0	$239	$0	$0	$483	$46	0%	$0	15%	$79	$608	$1
11.8	Main Power Transformers	$17,366	$0	$137	$0	$0	$17,503	$1,323	0%	$0	15%	$2,824	$21,651	$48
11.9	Electrical Foundations	$0	$158	$416	$0	$0	$575	$55	0%	$0	30%	$189	$818	$2
12	**INSTRUMENTATION & CONTROL**	$12,025	$2,213	$7,747	$0	$0	$21,985	$1,993	5%	$1,099	17%	$4,178	$29,256	$65
12.1	IGCC Control Equipment	w/4.1	$0	w/4.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.2	Combustion Turbine Control	w/6.1	$0	w/6.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.4	Other Major Component Control	$1,187	$0	$793	$0	$0	$1,980	$188	5%	$99	15%	$340	$2,607	$6
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.6	Control Boards,Panels & Racks	$273	$0	$174	$0	$0	$447	$43	5%	$22	20%	$102	$615	$1
12.7	Computer & Accessories	$6,332	$0	$202	$0	$0	$6,535	$600	5%	$327	10%	$746	$8,207	$18
12.8	Instrument Wiring & Tubing	$0	$2,213	$4,522	$0	$0	$6,735	$572	5%	$337	25%	$1,911	$9,554	$21
12.9	Other I & C Equipment	$4,233	$0	$2,055	$0	$0	$6,288	$592	5%	$314	15%	$1,079	$8,273	$18
13	**IMPROVEMENTS TO SITE**	$3,698	$2,178	$9,122	$0	$0	$14,999	$1,481	0%	$0	30%	$4,944	$21,424	$48
13.1	Site Preparation	$0	$116	$2,480	$0	$0	$2,596	$258	0%	$0	30%	$856	$3,710	$8
13.2	Site Improvements	$0	$2,062	$2,741	$0	$0	$4,803	$474	0%	$0	30%	$1,583	$6,860	$15
13.3	Site Facilities	$3,698	$0	$3,902	$0	$0	$7,600	$749	0%	$0	30%	$2,505	$10,854	$24
14	**BUILDINGS & STRUCTURES**	$0	$7,065	$7,948	$0	$0	$15,013	$1,365	0%	$0	16%	$2,693	$19,071	$43
14.1	Combustion Turbine Area	$0	$295	$167	$0	$0	$463	$40	0%	$0	20%	$101	$603	$1
14.2	Steam Turbine Building	$0	$2,294	$3,268	$0	$0	$5,563	$512	0%	$0	15%	$911	$6,986	$16
14.3	Administration Building	$0	$963	$699	$0	$0	$1,662	$148	0%	$0	15%	$271	$2,081	$5
14.4	Circulation Water Pumphouse	$0	$177	$94	$0	$0	$271	$24	0%	$0	15%	$44	$339	$1
14.5	Water Treatment Buildings	$0	$518	$506	$0	$0	$1,024	$92	0%	$0	15%	$167	$1,284	$3
14.6	Machine Shop	$0	$500	$342	$0	$0	$841	$75	0%	$0	15%	$137	$1,054	$2
14.7	Warehouse	$0	$806	$520	$0	$0	$1,326	$117	0%	$0	15%	$217	$1,660	$4
14.8	Other Buildings & Structures	$0	$473	$369	$0	$0	$842	$75	0%	$0	20%	$183	$1,100	$2
14.9	Waste Treating Building & Str.	$0	$1,037	$1,983	$0	$0	$3,021	$282	0%	$0	20%	$661	$3,964	$9
	TOTAL COST	$895,222	$77,962	$280,851	$0	$0	$1,254,035	$116,491	4%	$56,421	16%	$226,258	$1,653,205	$3,685

Table A5-5. O&M cost summary for the reference IGCC case.

Case	S3B Siemens 445 MW IGCC w/CO2	
Plant size	445.29 MW, net	
	Capacity factor	80%

Operating and Maintenance Labor

Operating Labor				
Operating Labor Rate	39.7	$/hour		
Operating Labor Burden Rate	30%	of base		
Labor Over Head Charge Rate	25%	of labor		
Administrative & Support Labor	25%	of burdened O&M labor		
Maintenance labor	0.914%	of TPC*		
Work force	**16**			
Skilled Operator	2			
Operator	10			
Foreman	1			
Lab Technician, etc	3			
			Annual Cost ($)	**Annual Unit Cost ($/MW)**
Fixed Operating Costs				
Operating Labor			$7,233,658	$16,244.824
Maintenance Labor			$16,030,126	$35,999.296
Administrative & Support Labor			$5,815,946	$13,061.030
Property Taxes & Insurance			$35,071,973	$78,762.094
TOTAL FIXED OPERATING COST			**$64,151,703**	**$144,067.243**

* (back calculated from NETL)

Variable Operating Costs

					Annual Cost ($)	**Annual Unit Cost ($/kW.h)**
Maintenance Material Cost	1.745%	of TPC*			**$ 30,599,792**	**$0.00981**
	Consumption					
	Initial	**per day**	**Unit Cost ($)**	**Initial Cost ($)**	**Annual Cost ($)**	**Annual Unit Cost ($/kW.h)**
Water (1000 gallons)	0.00	2,898	1.67	**$0**	**$1,413,279**	**$0.00045**
Chemicals				**$15,245,259**	**$3,379,909**	**$0.00108**
MU & WT Chem (lb)	0	17,265	0.27	$0	1,361,192	$0.00044
Carbon (Mercury Removal) (lb)	114,477	157	1.63	$186,598	74,639	$0.00002
COS Catalyst (m3)	0	0	3751.70	$0	0	$0.00000
Water Gas Shift Catalyst (ft3)	6,049	4.14	771.99	$4,670,088	934,018	$0.00030
Selexol Solution (Gal)	282,375	90	36.79	$10,388,574	964,958	$0.00031
SCR Catalyst (m3)	0	0	0	$0	0	$0.00000
Aqueous Amonia (ton)	0	0	0	$0	0	$0.00000
Claus Catalyst (ft3)	0	0.76	203.15	$0	45,102	$0.00001
Other				**$0**	**$0**	**$0.00000**
Supplemental Fuel (MMBtu)	0	0		$0	$0	$0.00000
Gases, N2 etc (100 scf)	0	0		$0	$0	$0.00000
L.P. Steam (1000 lb)	0	0		$0	$0	$0.00000
Waste Disposal				**$0**	**$70,956,811**	**$0.02274**
Spent Mercury Catalyst (lb)	0	157	0.65	$0	$29,841	$0.00001
Flyash (ton)	0	0	0.00	$0	$0	$0.00000
Slag (ton)	0	587	25.11	$0	$4,303,987	$0.00138
CO2 TS&M (1000 kg)	*0*	*10,371*	*22.00*	*$0*	*$66,622,983*	*$0.02135*
By-products and Emissions				**$0**	**$0**	**$0.00000**
Sulfur (ton)	0	50	0	$0	$0	$0.00000
TOTAL VARIABLE OPERATING COSTS				**$15,245,259**	**$106,349,791**	**$0.03408**
Fuel				**$0**	**$39,882,511**	**$0.01278**
Coal (ton)	0	6,958	19.63	$0	$39,882,511	$0.01278

Table A5-6. O&M cost summary for the Sour PSA IGCC case.

Case	Sour PSA with AGE-Claus		
Plant size	448.6294 MW, net		
		Capacity factor	80%

Operating and Maintenance Labor

Operating Labor				
Operating Labor Rate	39.7	$/hour		
Operating Labor Burden Rate	30%	of base		
Labor Over Head Charge Rate	25%	of labor		
Administrative & Support Labor	25%	of burdened O&M labor		
Maintenance labor	0.914%	of TPC*		
Work force	**16**			
Skilled Operator	2			
Operator	10			
Foreman	1			
Lab Technician, etc	3			
			Annual Cost ($)	**Annual Unit Cost ($/MW)**
Fixed Operating Costs				
Operating Labor			$7,233,658	$16,123.906
Maintenance Labor			$15,112,404	$33,685.722
Administrative & Support Labor			$5,586,515	$12,452.407
Property Taxes & Insurance			$33,064,107	$73,700.276
TOTAL FIXED OPERATING COST			**$60,996,683**	**$135,962.311**

* (back calculated from NETL)

Variable Operating Costs

					Annual Cost ($)	**Annual Unit Cost ($/kW.h)**
Maintenance Material Cost	1.745%	of TPC*			**$ 28,847,957**	**$0.00918**
	Consumption					
	Initial	**per day**	**Unit Cost ($)**	**Initial Cost ($)**	**Annual Cost ($)**	**Annual Unit Cost ($/kW.h)**
Water (1000 gallons)	0.00	2,938	1.67	**$0**	**$1,432,880**	**$0.00046**
Chemicals				**$13,943,500**	**$3,515,811**	**$0.00112**
MU & WT Chem (lb)	0	17,265	0.27	$0	$1,361,192	$0.00043
Carbon (Mercury Removal) (lb)	114,477	157	1.63	$186,598	$74,639	$0.00002
COS Catalyst (m3)	0	0	3751.70	$0	$0	$0.00000
Water Gas Shift Catalyst (ft3)	6,049	4.14	771.99	$4,670,088	$934,018	$0.00030
AGE (Flexorb) Solution (Gal)	-	-	-	$5,390,000	$808,500	$0.00026
SCR Catalyst (m3)	0	0	0	$0	$0	$0.00000
Aqueous Amonia (ton)	0	0	0	$0	$0	$0.00000
Sour PSA adsorbant (lb)	3,309,000	907	1.1172	$3,696,815	$295,745	$0.00009
Claus Catalyst (ft3)	0	0.70	203.15	$0	$41,717	$0.00001
Other				**$0**	**$0**	**$0.00000**
Supplemental Fuel (MMBtu)	0	0		$0	$0	$0.00000
Gases, N2 etc (100 scf)	0	0		$0	$0	$0.00000
L.P. Steam (1000 lb)	0	0		$0	$0	$0.00000
Waste Disposal				**$0**	**$71,412,952**	**$0.02271**
Spent Mercury Catalyst (lb)	0	157	0.65	$0	$29,841	$0.00001
Flyash (ton)	0	0	0.00	$0	$0	$0.00000
Slag (ton)	0	587	25.11	$0	$4,303,987	$0.00137
CO2 TS&M (1000 kg)	*0*	*10,442*	*22.00*	*$0*	*$67,079,124*	*$0.02134*
By-products and Emissions				**$0**	**$0**	**$0.00000**
Sulfur (ton)	0	50	0	$0	$0	$0.00000
TOTAL VARIABLE OPERATING COSTS				**$13,943,500**	**$105,209,600**	**$0.03346**
Fuel				**$0**	**$39,882,511**	**$0.01269**
Coal (ton)	0	6,958	19.63	$0	$39,882,511	$0.01269
Natural Gas (MMBtu-HHV)	0	0.00	5.17	$0	$0	$0.00000

Table A5-7. Cash flow analysis for the reference IGCC case, in $ x1000.

plant net output	445.29	MW
Construction period	5	year
estimated COE	157.01	mills/kW.h

end of	year 1	year 2	year 3	year 4	year 5	year 6	year 7	year 8	year 9	year 10	year 11	year 12	year 13	year 14	year 15
CO2 sell at gate						0	0	0	0	0	0	0	0	0	0
Operating revenues	$ -	$ -	$ -	$ -	$ -	$ 490,301	$ 505,010	$ 520,160	$ 535,765	$ 551,838	$ 568,393	$ 585,445	$ 603,008	$ 621,098	$ 639,731
Operating expenses															
Fixed	$ -	$ -	$ -	$ -	$ -	$ 74,369	$ 76,600	$ 78,899	$ 81,265	$ 83,703	$ 86,215	$ 88,801	$ 91,465	$ 94,209	$ 97,035
Variable	$ -	$ -	$ -	$ -	$ -	$ 123,289	$ 126,987	$ 130,797	$ 134,721	$ 138,762	$ 142,925	$ 147,213	$ 151,629	$ 156,178	$ 160,864
Fuel	$ -	$ -	$ -	$ -	$ -	$ 46,235	$ 47,622	$ 49,050	$ 50,522	$ 52,038	$ 53,599	$ 55,207	$ 56,863	$ 58,569	$ 60,326
CO2 taxe						$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating income	$ -	$ -	$ -	$ -	$ -	$ 246,408	$ 253,800	$ 261,414	$ 269,257	$ 277,334	$ 285,654	$ 294,224	$ 303,051	$ 312,142	$ 321,507
Interest Expense	$ -	$ -	$ -	$ -	$ -	$ 53,284	$ 50,906	$ 48,397	$ 45,751	$ 42,958	$ 40,013	$ 36,905	$ 33,626	$ 30,167	$ 26,518
Depreciation & Amortisation	$ -	$ -	$ -	$ -	$ -	$ 76,534	$ 147,334	$ 136,272	$ 126,067	$ 116,598	$ 107,862	$ 99,760	$ 92,290	$ 91,066	$ 91,045
Taxable Income	$ -	$ -	$ -	$ -	$ -	$ 116,590	$ 55,561	$ 76,745	$ 97,439	$ 117,778	$ 137,779	$ 157,559	$ 177,134	$ 190,909	$ 203,943
Income Taxes	$ -	$ -	$ -	$ -	$ -	$ 44,304	$ 21,113	$ 29,163	$ 37,027	$ 44,756	$ 52,356	$ 59,872	$ 67,311	$ 72,545	$ 77,498
Net Income	$ -	$ -	$ -	$ -	$ -	$ 72,286	$ 34,448	$ 47,582	$ 60,412	$ 73,023	$ 85,423	$ 97,687	$ 109,823	$ 118,364	$ 126,445
Cash form Operation	$ -	$ -	$ -	$ -	$ -	$ 246,408	$ 253,800	$ 261,414	$ 269,257	$ 277,334	$ 285,654	$ 294,224	$ 303,051	$ 312,142	$ 321,507
Income Taxes	$ -	$ -	$ -	$ -	$ -	$ 44,304	$ 21,113	$ 29,163	$ 37,027	$ 44,756	$ 52,356	$ 59,872	$ 67,311	$ 72,545	$ 77,498
Total Interest Expense	$ -	$ -	$ -	$ -	$ -	$ 53,284	$ 50,906	$ 48,397	$ 45,751	$ 42,958	$ 40,013	$ 36,905	$ 33,626	$ 30,167	$ 26,518
Total Principal Repayment	$ -	$ -	$ -	$ -	$ -	$ 43,233	$ 45,611	$ 48,119	$ 50,766	$ 53,558	$ 56,504	$ 59,611	$ 62,890	$ 66,349	$ 69,998
Operating Cash Flow	$ -	$ -	$ -	$ -	$ -	$ 105,587	$ 136,171	$ 135,735	$ 135,714	$ 136,062	$ 136,782	$ 137,835	$ 139,223	$ 143,080	$ 147,492
Capital Cost	$ 175,360	$ 545,018	$ 470,533	$ 389,977	$ 303,012	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ (175,360)	$ (545,018)	$ (470,533)	$ (389,977)	$ (303,012)	$ 105,587	$ 136,171	$ 135,735	$ 135,714	$ 136,062	$ 136,782	$ 137,835	$ 139,223	$ 143,080	$ 147,492
Loan Draws	$ 78,912	$ 245,258	$ 211,740	$ 175,490	$ 136,356	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ (96,448)	$ (299,760)	$ (258,793)	$ (214,488)	$ (166,657)	$ 105,587	$ 136,171	$ 135,735	$ 135,714	$ 136,062	$ 136,782	$ 137,835	$ 139,223	$ 143,080	$ 147,492
Equity Draws	$ 96,448	$ 299,760	$ 258,793	$ 214,488	$ 166,657	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ -	$ -	$ -	$ -	$ -	$ 105,587	$ 136,171	$ 135,735	$ 135,714	$ 136,062	$ 136,782	$ 137,835	$ 139,223	$ 143,080	$ 147,492

Internal rate of Return	**12.00%**
Net Present Value at discount rate	
8%	$ 581,969
10%	$ 231,286
12%	$ 0

Table A5-7. Cash flow analysis for the reference IGCC case, in $ x1000 (cont.).

end of	year 16	year 17	year 18	year 19	year 20	year 21	year 22	year 23	year 24	year 25
CO2 sell at gate	0	0	0	0	0	0	0	0	0	0
Operating revenues	$ 658,923	$ 678,691	$ 699,052	$ 720,023	$ 741,624	$ 763,873	$ 786,789	$ 810,392	$ 834,704	$ 859,745
Operating expenses										
Fixed	$ 99,946	$ 102,945	$ 106,033	$ 109,214	$ 112,490	$ 115,865	$ 119,341	$ 122,921	$ 126,609	$ 130,407
Variable	$ 165,690	$ 170,660	$ 175,780	$ 181,053	$ 186,485	$ 192,080	$ 197,842	$ 203,777	$ 209,891	$ 216,187
Fuel	$ 62,136	$ 64,000	$ 65,920	$ 67,897	$ 69,934	$ 72,032	$ 74,193	$ 76,419	$ 78,712	$ 81,073
CO2 taxe	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating income	$ 331,152	$ 341,086	$ 351,319	$ 361,859	$ 372,714	$ 383,896	$ 395,413	$ 407,275	$ 419,493	$ 432,078
Interest Expense	$ 22,668	$ 18,607	$ 14,322	$ 9,801	$ 5,032	$ -	$ -	$ -	$ -	$ -
Depreciation & Amortisation	$ 91,066	$ 91,045	$ 91,066	$ 91,045	$ 91,066	$ 91,045	$ 91,066	$ 91,045	$ 91,066	$ 91,045
Taxable Income	$ 217,418	$ 231,434	$ 245,932	$ 261,012	$ 276,617	$ 292,850	$ 304,347	$ 316,230	$ 328,428	$ 341,033
Income Taxes	$ 82,619	$ 87,945	$ 93,454	$ 99,185	$ 105,114	$ 111,283	$ 115,652	$ 120,167	$ 124,802	$ 129,592
Net Income	$ 134,799	$ 143,489	$ 152,478	$ 161,828	$ 171,503	$ 181,567	$ 188,695	$ 196,062	$ 203,625	$ 211,440
Cash form Operation	$ 331,152	$ 341,086	$ 351,319	$ 361,859	$ 372,714	$ 383,896	$ 395,413	$ 407,275	$ 419,493	$ 432,078
Income Taxes	$ 82,619	$ 87,945	$ 93,454	$ 99,185	$ 105,114	$ 111,283	$ 115,652	$ 120,167	$ 124,802	$ 129,592
Total Interest Expense	$ 22,668	$ 18,607	$ 14,322	$ 9,801	$ 5,032	$ -	$ -	$ -	$ -	$ -
Total Principal Repayment	$ 73,848	$ 77,910	$ 82,195	$ 86,715	$ 91,485	$ -	$ -	$ -	$ -	$ -
Operating Cash Flow	$ 152,017	$ 156,625	$ 161,348	$ 166,157	$ 171,083	$ 272,613	$ 279,761	$ 287,108	$ 294,691	$ 302,486
Capital Cost	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ 152,017	$ 156,625	$ 161,348	$ 166,157	$ 171,083	$ 272,613	$ 279,761	$ 287,108	$ 294,691	$ 302,486
Loan Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ 152,017	$ 156,625	$ 161,348	$ 166,157	$ 171,083	$ 272,613	$ 279,761	$ 287,108	$ 294,691	$ 302,486
Equity Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ 152,017	$ 156,625	$ 161,348	$ 166,157	$ 171,083	$ 272,613	$ 279,761	$ 287,108	$ 294,691	$ 302,486

end of	year 26	year 27	year 28	year 29	year 30	year 31	year 32	year 33	year 34	year 35
CO2 sell at gate	0	0	0	0	0	0	0	0	0	0
Operating revenues	$ 885,538	$ 912,104	$ 939,467	$ 967,651	$ 996,680	$ 1,026,581	$ 1,057,378	$ 1,089,100	$ 1,121,773	$ 1,155,426
Operating expenses										
Fixed	$ 134,319	$ 138,349	$ 142,499	$ 146,774	$ 151,178	$ 155,713	$ 160,384	$ 165,196	$ 170,152	$ 175,256
Variable	$ 222,673	$ 229,353	$ 236,234	$ 243,321	$ 250,620	$ 258,139	$ 265,883	$ 273,860	$ 282,075	$ 290,538
Fuel	$ 83,505	$ 86,010	$ 88,591	$ 91,248	$ 93,986	$ 96,805	$ 99,709	$ 102,701	$ 105,782	$ 108,955
CO2 taxe	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating income	$ 445,040	$ 458,392	$ 472,143	$ 486,308	$ 500,897	$ 515,924	$ 531,401	$ 547,343	$ 563,764	$ 580,677
Interest Expense	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Depreciation & Amortisation	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Taxable Income	$ 445,040	$ 458,392	$ 472,143	$ 486,308	$ 500,897	$ 515,924	$ 531,401	$ 547,343	$ 563,764	$ 580,677
Income Taxes	$ 169,115	$ 174,189	$ 179,414	$ 184,797	$ 190,341	$ 196,051	$ 201,933	$ 207,991	$ 214,230	$ 220,657
Net Income	$ 275,925	$ 284,203	$ 292,729	$ 301,511	$ 310,556	$ 319,873	$ 329,469	$ 339,353	$ 349,534	$ 360,020
Cash form Operation	$ 445,040	$ 458,392	$ 472,143	$ 486,308	$ 500,897	$ 515,924	$ 531,401	$ 547,343	$ 563,764	$ 580,677
Income Taxes	$ 169,115	$ 174,189	$ 179,414	$ 184,797	$ 190,341	$ 196,051	$ 201,933	$ 207,991	$ 214,230	$ 220,657
Total Interest Expense	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Total Principal Repayment	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating Cash Flow	$ 275,925	$ 284,203	$ 292,729	$ 301,511	$ 310,556	$ 319,873	$ 329,469	$ 339,353	$ 349,534	$ 360,020
Capital Cost	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ 275,925	$ 284,203	$ 292,729	$ 301,511	$ 310,556	$ 319,873	$ 329,469	$ 339,353	$ 349,534	$ 360,020
Loan Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ 275,925	$ 284,203	$ 292,729	$ 301,511	$ 310,556	$ 319,873	$ 329,469	$ 339,353	$ 349,534	$ 360,020
Equity Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ 275,925	$ 284,203	$ 292,729	$ 301,511	$ 310,556	$ 319,873	$ 329,469	$ 339,353	$ 349,534	$ 360,020

Table A5-8. Cash flow analysis for the Sour PSA IGCC case, in $ x1000.

plant net output	448.63	MW
Construction period	5	year
estimated COE	149.78	mills/kW.h

end of	year 1	year 2	year 3	year 4	year 5	year 6	year 7	year 8	year 9	year 10	year 11	year 12	year 13	year 14	year 15
CO2 sell at gate						$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating revenues	$ -	$ -	$ -	$ -	$ -	$ 471,231	$ 485,368	$ 499,929	$ 514,927	$ 530,375	$ 546,286	$ 562,675	$ 579,555	$ 596,941	$ 614,850
Operating expenses															
Fixed	$ -	$ -	$ -	$ -	$ -	$ 70,712	$ 72,833	$ 75,018	$ 77,269	$ 79,587	$ 81,974	$ 84,434	$ 86,967	$ 89,576	$ 92,263
Variable	$ -	$ -	$ -	$ -	$ -	$ 121,967	$ 125,626	$ 129,395	$ 133,276	$ 137,275	$ 141,393	$ 145,635	$ 150,004	$ 154,504	$ 159,139
Fuel	$ -	$ -	$ -	$ -	$ -	$ 46,235	$ 47,622	$ 49,050	$ 50,522	$ 52,038	$ 53,599	$ 55,207	$ 56,863	$ 58,569	$ 60,326
CO2 taxe						$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating income	$ -	$ -	$ -	$ -	$ -	$ 232,318	$ 239,287	$ 246,466	$ 253,860	$ 261,476	$ 269,320	$ 277,399	$ 285,721	$ 294,293	$ 303,122
Interest Expense	$ -	$ -	$ -	$ -	$ -	$ 50,243	$ 48,001	$ 45,636	$ 43,140	$ 40,507	$ 37,730	$ 34,799	$ 31,708	$ 28,446	$ 25,005
Depreciation & Amortisation	$ -	$ -	$ -	$ -	$ -	$ 72,153.90	$ 138,901	$ 128,472	$ 118,852	$ 109,924	$ 101,689	$ 94,050	$ 87,008	$ 85,854	$ 85,834
Taxable Income	$ -	$ -	$ -	$ -	$ -	$ 109,920	$ 52,385	$ 72,358	$ 91,868	$ 111,044	$ 129,901	$ 148,550	$ 167,006	$ 179,993	$ 192,282
Income Taxes	$ -	$ -	$ -	$ -	$ -	$ 41,770	$ 19,906	$ 27,496	$ 34,910	$ 42,197	$ 49,362	$ 56,449	$ 63,462	$ 68,397	$ 73,067
Net Income	$ -	$ -	$ -	$ -	$ -	$ 68,151	$ 32,479	$ 44,862	$ 56,958	$ 68,847	$ 80,539	$ 92,101	$ 103,543	$ 111,596	$ 119,215
Cash form Operation	$ -	$ -	$ -	$ -	$ -	$ 232,318	$ 239,287	$ 246,466	$ 253,860	$ 261,476	$ 269,320	$ 277,399	$ 285,721	$ 294,293	$ 303,122
Income Taxes	$ -	$ -	$ -	$ -	$ -	$ 41,770	$ 19,906	$ 27,496	$ 34,910	$ 42,197	$ 49,362	$ 56,449	$ 63,462	$ 68,397	$ 73,067
Total Interest Expense	$ -	$ -	$ -	$ -	$ -	$ 50,243	$ 48,001	$ 45,636	$ 43,140	$ 40,507	$ 37,730	$ 34,799	$ 31,708	$ 28,446	$ 25,005
Total Principal Repayment	$ -	$ -	$ -	$ -	$ -	$ 40,766	$ 43,008	$ 45,374	$ 47,869	$ 50,502	$ 53,280	$ 56,210	$ 59,302	$ 62,563	$ 66,004
Operating Cash Flow	$ -	$ -	$ -	$ -	$ -	$ 99,538	$ 128,371	$ 127,960	$ 127,940	$ 128,269	$ 128,948	$ 129,941	$ 131,250	$ 134,886	$ 139,045
Capital Cost	$ 165,321	$ 513,816	$ 443,595	$ 367,651	$ 285,665	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ (165,321)	$ (513,816)	$ (443,595)	$ (367,651)	$ (285,665)	$ 99,538	$ 128,371	$ 127,960	$ 127,940	$ 128,269	$ 128,948	$ 129,941	$ 131,250	$ 134,886	$ 139,045
Loan Draws	$ 74,394	$ 231,217	$ 199,618	$ 165,443	$ 128,549	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ (90,926)	$ (282,599)	$ (243,977)	$ (202,208)	$ (157,116)	$ 99,538	$ 128,371	$ 127,960	$ 127,940	$ 128,269	$ 128,948	$ 129,941	$ 131,250	$ 134,886	$ 139,045
Equity Draws	$ 90,926	$ 282,599	$ 243,977	$ 202,208	$ 157,116	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ -	$ -	$ -	$ -	$ -	$ 99,538	$ 128,371	$ 127,960	$ 127,940	$ 128,269	$ 128,948	$ 129,941	$ 131,250	$ 134,886	$ 139,045

Internal rate of Return	**12.00%**
Net Present Value at discount rate	
8%	$ 548,668
10%	$ 218,051
12%	$ 0

Table A5-8. Cash flow analysis for the Sour PSA IGCC case, in $ x1000 (cont.).

end of	year 16	year 17	year 18	year 19	year 20	year 21	year 22	year 23	year 24	year 25
CO2 sell at gate	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating revenues	$ 633,295	$ 652,294	$ 671,863	$ 692,019	$ 712,779	$ 734,163	$ 756,187	$ 778,873	$ 802,239	$ 826,306
Operating expenses										
Fixed	$ 95,031	$ 97,882	$ 100,818	$ 103,843	$ 106,958	$ 110,167	$ 113,472	$ 116,876	$ 120,382	$ 123,994
Variable	$ 163,913	$ 168,831	$ 173,895	$ 179,112	$ 184,486	$ 190,020	$ 195,721	$ 201,592	$ 207,640	$ 213,869
Fuel	$ 62,136	$ 64,000	$ 65,920	$ 67,897	$ 69,934	$ 72,032	$ 74,193	$ 76,419	$ 78,712	$ 81,073
CO2 taxe	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating income	$ 312,215	$ 321,582	$ 331,229	$ 341,166	$ 351,401	$ 361,943	$ 372,802	$ 383,986	$ 395,505	$ 407,370
Interest Expense	$ 21,375	$ 17,545	$ 13,505	$ 9,242	$ 4,745	$ -	$ -	$ -	$ -	$ -
Depreciation & Amortisation	$ 85,854	$ 85,834	$ 85,854	$ 85,834	$ 85,854	$ 85,834	$ 85,854	$ 85,834	$ 85,854	$ 85,834
Taxable Income	$ 204,987	$ 218,203	$ 231,871	$ 246,090	$ 260,803	$ 276,109	$ 286,948	$ 298,151	$ 309,652	$ 321,536
Income Taxes	$ 77,895	$ 82,917	$ 88,111	$ 93,514	$ 99,105	$ 104,921	$ 109,040	$ 113,298	$ 117,668	$ 122,184
Net Income	$ 127,092	$ 135,286	$ 143,760	$ 152,576	$ 161,698	$ 171,188	$ 177,908	$ 184,854	$ 191,984	$ 199,352
Cash form Operation	$ 312,215	$ 321,582	$ 331,229	$ 341,166	$ 351,401	$ 361,943	$ 372,802	$ 383,986	$ 395,505	$ 407,370
Income Taxes	$ 77,895	$ 82,917	$ 88,111	$ 93,514	$ 99,105	$ 104,921	$ 109,040	$ 113,298	$ 117,668	$ 122,184
Total Interest Expense	$ 21,375	$ 17,545	$ 13,505	$ 9,242	$ 4,745	$ -	$ -	$ -	$ -	$ -
Total Principal Repayment	$ 69,635	$ 73,465	$ 77,505	$ 81,768	$ 86,265	$ -	$ -	$ -	$ -	$ -
Operating Cash Flow	$ 143,311	$ 147,655	$ 152,109	$ 156,642	$ 161,286	$ 257,022	$ 263,761	$ 270,688	$ 277,838	$ 285,187
Capital Cost	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ 143,311	$ 147,655	$ 152,109	$ 156,642	$ 161,286	$ 257,022	$ 263,761	$ 270,688	$ 277,838	$ 285,187
Loan Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ 143,311	$ 147,655	$ 152,109	$ 156,642	$ 161,286	$ 257,022	$ 263,761	$ 270,688	$ 277,838	$ 285,187
Equity Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ 143,311	$ 147,655	$ 152,109	$ 156,642	$ 161,286	$ 257,022	$ 263,761	$ 270,688	$ 277,838	$ 285,187

end of	year 26	year 27	year 28	year 29	year 30	year 31	year 32	year 33	year 34	year 35
CO2 sell at gate	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating revenues	$ 851,096	$ 876,629	$ 902,927	$ 930,015	$ 957,916	$ 986,653	$ 1,016,253	$ 1,046,740	$ 1,078,143	$ 1,110,487
Operating expenses										
Fixed	$ 127,714	$ 131,545	$ 135,491	$ 139,556	$ 143,743	$ 148,055	$ 152,497	$ 157,072	$ 161,784	$ 166,637
Variable	$ 220,286	$ 226,894	$ 233,701	$ 240,712	$ 247,933	$ 255,371	$ 263,032	$ 270,923	$ 279,051	$ 287,423
Fuel	$ 83,505	$ 86,010	$ 88,591	$ 91,248	$ 93,986	$ 96,805	$ 99,709	$ 102,701	$ 105,782	$ 108,955
CO2 taxe	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating income	$ 419,592	$ 432,179	$ 445,145	$ 458,499	$ 472,254	$ 486,422	$ 501,014	$ 516,045	$ 531,526	$ 547,472
Interest Expense	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Depreciation & Amortisation	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Taxable Income	$ 419,592	$ 432,179	$ 445,145	$ 458,499	$ 472,254	$ 486,422	$ 501,014	$ 516,045	$ 531,526	$ 547,472
Income Taxes	$ 159,445	$ 164,228	$ 169,155	$ 174,230	$ 179,456	$ 184,840	$ 190,385	$ 196,097	$ 201,980	$ 208,039
Net Income	$ 260,147	$ 267,951	$ 275,990	$ 284,269	$ 292,797	$ 301,581	$ 310,629	$ 319,948	$ 329,546	$ 339,432
Cash form Operation	$ 419,592	$ 432,179	$ 445,145	$ 458,499	$ 472,254	$ 486,422	$ 501,014	$ 516,045	$ 531,526	$ 547,472
Income Taxes	$ 159,445	$ 164,228	$ 169,155	$ 174,230	$ 179,456	$ 184,840	$ 190,385	$ 196,097	$ 201,980	$ 208,039
Total Interest Expense	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Total Principal Repayment	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Operating Cash Flow	$ 260,147	$ 267,951	$ 275,990	$ 284,269	$ 292,797	$ 301,581	$ 310,629	$ 319,948	$ 329,546	$ 339,432
Capital Cost	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ 260,147	$ 267,951	$ 275,990	$ 284,269	$ 292,797	$ 301,581	$ 310,629	$ 319,948	$ 329,546	$ 339,432
Loan Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ 260,147	$ 267,951	$ 275,990	$ 284,269	$ 292,797	$ 301,581	$ 310,629	$ 319,948	$ 329,546	$ 339,432
Equity Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ 260,147	$ 267,951	$ 275,990	$ 284,269	$ 292,797	$ 301,581	$ 310,629	$ 319,948	$ 329,546	$ 339,432

Project: DE-FE0013363

Table A5-9. Stream mass balance for the reference coal-to-methanol case.

stream number	100	101	102	103	104	105	106	201	202	203	204	205
Description	Wet coal	Air	Feed air	Dry coal	Exhaust	LP Recycle gas	HP Recycle gas	ASU air	Hot O2	Claus O2	Coal Drying N2	Casifier CO2
Mole Flow (Vap/Liq) lbmol/h												
Ar	-	-	-		269	95	95	858	604	3	-	-
CH4	-	-	-		78,959	27,743	27,743	-	-	-	-	-
CO	-	-	-		297	104	104	-	-	-	-	24
CO_2	-	-	-		-	-	-	30	-	-	-	3,582
H2S	-	-	-		2,952	1,037	1,037	-	-	-	-	-
COS	-	-	-		-	-	-	-	-	-	-	-
HCN	-	-	-		14,736	5,178	5,178	-	-	-	-	-
H2	-	-	-	-	-	-	-	-	-	-	-	6
H2O	-	-	-	-	12,143	4,262	4,262	915	-	-	-	-
N2	-	-	-	-	22,284	7,822	7,822	71,692	403	2	71,240	-
NH3	-	-	-	-	-	-	-	-	-	-	-	-
O2	-	-	-	-	-	-	-	19,232	19,129	97	-	-
SO2	-	-	-	-	3	1	1	-	-	-	-	-
Mass Flow (Sol) lb/hr												
Coal	1,041,742			822,641								
Ash												
Slag												
Sulfur												
Mole Flow Rate (lbmol/hr)	-	1,500	1,500	-	131,644	46,242	46,242	92,727	20,136	102	71,240	3,612
Mass Flow Rate (lb./hr)	1,041,742	-	-	822,641	2,627,811	922,994	922,994	2,675,529	647,511	3,293	1,995,396	158,321
Temperature (F)	59.00	59.00	59.00	157.00	164.60	164.60	180.15	59.00	300.00	70.00	79.00	269.76
Pressure (PSIA)	14.70	14.70	17.00	15.00	15.00	15.00	16.00	14.70	710.88	14.70	18.00	768.89

Table A5-9. Stream mass balance for the reference coal-to-methanol case (cont.).

stream number	209	212	213	214	215	303	304	305	306	307	308	309
Description	Raw syngas	Cold quench water	Hot quench water	Slag	Dry solids	Syngas to shift	Bypass syngas	Hot syngas to shift	Hot 1st shift syngas	Warm 1st shift syngas	Cold1st shift syngas	Hot 2nd shift syngas
Mole Flow (Vap/Liq) lbmol/h												
Ar	1,017	-	-	-	-	369	236	369	369	369	369	369
CH4	6	-	-	-	-	3	1	3	3	3	3	3
CO	71,520	-	-	-	-	25,914	16,558	25,914	7,727	7,727	7,727	2,869
CO_2	7,405	-	-	-	-	2,699	1,725	2,699	20,897	20,897	20,897	25,755
H2S	368	-	-	-	-	134	86	134	145	145	145	145
COS	33	-	-	-	-	12	8	12	1	1	1	-
HCN	1	-	-	-	-	1	1	1	-	-	-	-
H2	26,615	-	-	-	-	9,646	6,164	9,646	27,834	27,834	27,834	32,692
H2O	96,734	39,597	39,597	-	-	42,207	26,969	42,207	24,008	24,008	24,008	19,149
N2	1,122	-	-	-	-	407	260	407	407	407	407	407
NH3	10	-	-	-	-	8	5	8	9	9	9	9
O2	-	-	-	-	-	-	-	-	-	-	-	-
SO2	-	-	-	-	-	-	-	-	-	-	-	-
Mass Flow (Sol) lb/hr												
Coal												
Ash					17,064							
Slag				54,688								
Sulfur												
Mole Flow Rate (lbmol/hr)	204,832	39,597	39,597	-	-	81,399	52,012	81,399	81,400	81,400	81,400	81,398
Mass Flow Rate (lb./hr)	4,212,423	713,354	713,354	-	-	1,656,051	1,058,195	1,656,051	1,656,073	1,656,073	1,656,073	1,656,001
Temperature (F)	603.32	59.00	392.00	2,650.00	612.32	415.31	415.31	530.00	938.81	600.00	495.16	606.40
Pressure (PSIA)	605.34	700.00	685.00	650.00	630.00	605.34	605.34	600.34	593.09	584.39	575.69	567.69

Table A5-9. Stream mass balance for the reference coal-to-methanol case (cont.).

stream number	311	312	313	314	315	316	317	318	401	402	403	404
Description	Scrubber water	Scrubber hot vapor	Scrubber cold vapor	Scrubber sour gas	LP recycle water	HP recycle water	Bottoms to WWT	Recycle syngas	Shifted syngas	Shifted syngas	Shifted syngas	Shifted syngas
Mole Flow (Vap/Liq) lbmol/h												
Ar	-	-	-	-	-	-	-	413	369	369	369	604
CH4	-	-	-	-	-	-	-	3	3	3	3	4
CO	6	6	6	6	-	-	-	29,041	2,869	2,869	2,869	19,428
CO_2	3	3	3	-	-	-	-	3,024	25,755	25,755	25,755	27,480
H2S	1	1	1	-	-	-	-	150	145	145	145	231
COS	-	-	-	-	-	-	-	14	-	-	-	8
HCN	-	-	-	-	-	-	-	1	-	-	-	1
H2	3	3	3	3	-	-	-	10,811	32,692	32,692	32,692	38,855
H2O	40,422	6,123	6,123	1	6,117	6,117	34,299	47,300	19,149	19,149	19,149	46,118
N2	-	-	-	-	-	-	-	456	407	407	407	668
NH3	1	-	-	-	-	-	-	9	9	9	9	14
O2	-	-	-	-	-	-	-	-	-	-	-	-
SO2	-	-	-	-	-	-	-	-	-	-	-	-
Mass Flow (Sol) lb/hr												
Coal												
Ash	5,784											
Slag												
Sulfur												
Mole Flow Rate (lbmol/hr)	40,436	6,136	6,136	10	6,117	6,117	34,299	91,222	81,398	81,398	81,398	133,410
Mass Flow Rate (lb./hr)	728,565	110,638	110,638	180	110,202	110,202	617,916	1,855,906	1,656,023	1,656,023	1,656,023	2,714,208
Temperature (F)	415.31	281.93	130.00	130.00	130.00	130.96	181,93	415.31	536.00	489.29	409.06	406.45
Pressure (PSIA)	605.34	50.76	50.76	50.76	50.76	550.00	50.76	605.34	562.69	557.69	552.69	552.69

Table A5-9. Stream mass balance for the reference coal-to-methanol case (cont.).

stream number	405	406	407	408	409	410	411	412	413	414	415	416
Description	Shifted syngas	Shifted syngas	Shifted syngas	Shifted syngas	Shifted syngas	KO water	Sour water	Sour flash gas	Sour stripper feed	Sour stipper gas	Sour gas	Sour stripper bottoms
Mole Flow (Vap/Liq) lbmol/h												
Ar	604	604	604	604	604	-	-	-	-	-	-	-
CH4	4	4	4	4	4	-	-	-	-	-	-	-
CO	19,428	19,427	19,427	19,426	19,424	2	1	-	1	1	1	-
CO_2	27,480	27,470	27,470	27,450	27,404	26	37	-	37	37	37	-
H2S	231	230	230	230	229	1	1	-	1	1	1	1
COS	8	8	8	8	8	-	-	-	-	-	-	-
HCN	1	1	1	1	-	-	-	-	-	-	-	-
H2	38,855	38,853	38,853	38,850	38,846	6	3	-	3	3	3	-
H2O	46,118	32,555	32,555	15,497	202	30,618	15,286	-	15,286	139	139	15,147
N2	668	668	668	667	667	-	-	-	-	-	-	-
NH3	14	13	13	10	-	1	-	-	-	-	-	-
O2	-	-	-	-	-	-	-	-	-	-	-	-
SO2	-	-	-	-	-	-	-	-	-	-	-	-
Mass Flow (Sol) lb/hr												
Coal												
Ash												
Slag												
Sulfur												
Mole Flow Rate (lbmol/hr)	133,410	119,833	119,833	102,746	87,388	30,654	15,329	-	15,329	182	182	15,148
Mass Flow Rate (lb./hr)	2,714,208	2,469,387	2,469,387	2,161,088	1,883,217	552,861	277,090	-	277,090	4,206	4,206	272,905
Temperature (F)	366.96	350.00	305.78	110.63	110.63	327.11	17,895.00	178.95	178.95	276.97	276.97	302.02

Table A5-9. Stream mass balance for the reference coal-to-methanol case (cont.).

stream number	418	419	500	501	502	503	504	505	506	601	602
Description	Cold condesate	Hot condensate to scrubber	Cold weet syngas	Hot sweet syngas	LP CO2	HP CO2	KO water	Acid gas to claus	CO2 vent	Sulfur product	TG to coal drying
Mole Flow (Vap/Liq) lbmol/h											
Ar	-	-	604	604	-	-	-	-	-	-	3
CH4	-	-	4	4	-	-	-	-	-	-	-
CO	2	2	19,262	19,262	162	138	-	1	24	-	42
CO_2	30	35	2,217	2,217	24,604	21,022	-	583	3,582	-	595
H2S	1	1	-	-	-	-	-	229	-	-	3
COS	-	-	-	-	-	-	-	8	-	-	1
HCN	-	-	-	-	-	-	-	-	-	-	-
H2	6	6	38,800	38,800	46	39	-	1	6	-	15
H2O	51,885	51,890	192	192	-	-	-	10	-	-	375
N2	-	-	667	667	-	-	-	-	-	-	3
NH3	4	10	-	-	-	-	-	-	-	-	-
O2	-	-	-	-	-	-	-	-	-	-	-
SO2	-	-	-	-	-	-	-	-	-	-	1
Mass Flow (Sol) lb/hr											
Coal											
Ash											
Slag											
Sulfur										11,576	
Mole Flow Rate (lbmol/hr)	51,928	51,945	61,745	61,745	24,812	21,199	-	832	3,612	45	1,037
Mass Flow Rate (lb./hr)	936,200	936,658	761,613	761,613	1,087,450	929,129	-	34,149	158,321	11,576	34,569
Temperature (F)	299.48	392.00	68.00	125.00	57.20	196.00	100.00	86.00	269.76	387.64	320.00
Pressure (PSIA)	768.89	761.64	495.83	493.83	14.50	2,214.70	15.00	29.01	768.89	20.01	20.01

Table A5-9. Stream mass balance for the reference coal-to-methanol case (cont.).

stream number	701	702	703	704	705	707	708	709	710	711
Description	HP sweet syngas	cold mixed feed	Warm mixed feed	Hot mixed feed	Stage 1 product	Stage 2 cooled feed	Hot stage 2 product	Warm stage 2 product	Cold stage 2 product	Flash gas
Mole Flow (Vap/Liq) lbmol/h										
Ar	604	13,164	13,164	13,164	13,164	13,164	13,164	13,164	13,164	13,084
N2	667	15,098	15,098	15,098	15,098	15,098	15,098	15,098	15,098	15,032
CO	19,262	32,651	32,651	32,651	20,021	20,021	13,978	13,978	13,978	13,948
CO2	2,217	23,376	23,376	23,376	23,083	23,083	23,259	23,259	23,259	22,041
H2	38,800	64,754	64,754	64,754	38,616	38,616	27,058	27,058	27,058	27,034
H2O	192	202	202	202	500	500	330	330	330	11
CH4	4	77	77	77	77	77	77	77	77	76
CH3OH	-	1,775	1,775	1,775	14,687	14,687	20,544	20,544	20,544	1,205
C2H6	-	23	23	23	26	26	30	30	30	23
C3H8	-	-	-	-	1	1	2	2	2	-
Mole Flow Rate (lbmol/hr)	61,745	151,119	151,119	151,119	125,274	125,274	113,540	113,540	113,540	92,454
Mass Flow Rate (lb./hr)	761,613	3,085,008	3,085,008	3,085,008	3,084,930	3,084,930	3,084,881	3,084,881	3,084,881	2,399,569
Temperature (F)	228.60	173.42	437.73	400.00	475.00	400.00	430.00	230.00	130.00	130.00

Table A5-9. Stream mass balance for the reference coal-to-methanol case (cont.).

stream number	712	713	714	715	716	718	724	726	727	728	730
Description	LP recycle gas	HP recycle gas	Raw methanol	PSA feed	Raw methanol	PSA product	Methanol flash gas	purge gas	Purge gas to boiler	Purge gas to coal drying	Stack
Mole Flow (Vap/Liq) lbmol/h											
Ar	12,561	12,561	80	523	1	-	79	523	428	95	453
N2	14,431	14,431	66	601	1	-	65	601	492	109	2,601
CO	13,389	13,389	29	558	-	-	29	558	456	102	-
CO2	21,159	21,159	1,219	881	221	-	998	881	721	160	1,242
H2	25,954	25,954	23	1,082	-	703	23	379	310	69	-
H2O	10	10	319	1	317	-	1	1	1	-	440
CH4	73	73	1	3	-	-	1	3	3	1	-
CH3OH	1,775	1,775	18,694	74	18,429	-	266	74	61	14	-
C2H6	23	23	7	1	-	-	3	1	1	-	-
C3H8	23	23	2	-	-	-	-	-	-	-	-
Mole Flow Rate (lbmol/hr)	89,396	89,396	20,440	3,724	18,969	703	1,465	3,021	2,471	550	4,737
Mass Flow Rate (lb./hr)	2,324,343	2,324,343	664,627	96,803	606,007	1,417	58,388	95,385	78,028	17,357	153,571
Temperature (F)	130.00	141.50	130.00	130.00	117.46	70.00	117.46	70.00	70.00	70.00	246.00
Pressure (PSIA)	717.00	755.00	717.00	717.00	40.00	18.00	40.00	18.00	18.00	18.00	15.00

Table A5-10. Stream mass balance for the Sour PSA coal-to-methanol case.

stream number	100	101	102	103	104	105	106	201	202	203	204	205
Description	Wet coal	Air	Feed air	Dry coal	Exhaust	LP Recycle gas	HP Recycle gas	ASU air	Hot O2	Claus O2	Coal Drying N2	Casifier CO2
Mole Flow (Vap/Liq) lbmol/h												
Ar	-	-	-		269	95	95	858	604	3	-	-
CH4	-	-	-		78,959	27,743	27,743	-	-	-	-	-
CO	-	-	-		297	104	104	-	-	-	-	24
CO_2	-	-	-		-	-	-	30	-	-	-	3,582
H2S	-	-	-		2,952	1,037	1,037	-	-	-	-	-
COS	-	-	-		-	-	-	-	-	-	-	-
HCN	-	-	-		14,736	5,178	5,178	-	-	-	-	-
H2	-	-	-	-	-	-	-	-	-	-	-	6
H2O	-	-	-	-	12,143	4,262	4,262	915	-	-	-	-
N2	-	-	-	-	22,284	7,822	7,822	71,692	403	2	71,240	-
NH3	-	-	-	-	-	-	-	-	-	-	-	-
O2	-	-	-	-	-	-	-	19,232	19,129	97	-	-
SO2	-	-	-	-	3	1	1	-	-	-	-	-
Mass Flow (Sol) lb/hr												
Coal	1,041,742			822,641								
Ash												
Slag												
Sulfur												
Mole Flow Rate (lbmol/hr)	-	1,500	1,500	-	131,644	46,242	46,242	92,727	20,136	102	71,240	3,612
Mass Flow Rate (lb./hr)	1,041,742	-	-	822,641	2,627,811	922,994	922,994	2,675,529	647,511	3,293	1,995,396	158,321
Temperature (F)	59.00	59.00	59.00	157.00	164.60	164.60	180.15	59.00	300.00	70.00	79.00	269.76
Pressure (PSIA)	14.70	14.70	17.00	15.00	15.00	15.00	16.00	14.70	710.88	14.70	18.00	768.89

Table A5-10. Stream mass balance for the Sour PSA coal-to-methanol case (cont.).

stream number	209	212	213	214	215	303	304	305	306	307	308	309
Description	Raw syngas	Cold quench water	Hot quench water	Slag	Dry solids	Syngas to shift	Bypass syngas	Hot syngas to shift	Hot 1st shift syngas	Warm 1st shift syngas	Cold1st shift syngas	Hot 2nd shift syngas
Mole Flow (Vap/Liq) lbmol/h												
Ar	1,017	-	-	-	-	362	242	362	362	362	362	362
CH4	6	-	-	-	-	2	2	2	2	2	2	2
CO	71,520	-	-	-	-	25,467	17,006	25,467	7,594	7,594	7,594	2,820
CO_2	7,405	-	-	-	-	2,652	1,771	2,652	20,536	20,536	20,536	25,311
H2S	368	-	-	-	-	132	88	132	142	142	142	143
COS	33	-	-	-	-	12	8	12	1	1	1	-
HCN	1	-	-	-	-	1	1	1	-	-	-	-
H2	26,615	-	-	-	-	9,480	6,330	9,480	27,354	27,354	27,354	32,127
H2O	96,734	39,597	39,597	-	-	41,478	27,698	41,478	23,592	23,592	23,592	18,818
N2	1,122	-	-	-	-	400	267	400	400	400	400	400
NH3	10	-	-	-	-	8	5	8	9	9	9	9
O2	-	-	-	-	-	-	-	-	-	-	-	-
SO2	-	-	-	-	-	-	-	-	-	-	-	-
Mass Flow (Sol) lb/hr												
Coal												
Ash					17,064							
Slag				54,688								
Sulfur												
Mole Flow Rate (lbmol/hr)	204,832	39,597	39,597	-	-	79,993	53,418	79,993	79,993	79,993	79,993	79,993
Mass Flow Rate (lb./hr)	4,212,423	713,354	713,354	-	-	1,627,457	1,086,781	1,627,457	1,627,455	1,627,455	1,627,455	1,627,457
Temperature (F)	603.32	59.00	392.00	2,650.00	612.32	415.31	415.31	530.00	938.81	600.00	495.16	606.40

Table A5-10. Stream mass balance for the Sour PSA coal-to-methanol case (cont.).

stream number	311	312	313	314	315	316	317	318	401	402	403	404
Description	Scrubber water	Scrubber hot vapor	Scrubber cold vapor	Scrubber sour gas	LP recycle water	HP recycle water	Bottoms to WWT	Recycle syngas	Shifted syngas	Shifted syngas	Shifted syngas	Shifted syngas
Mole Flow (Vap/Liq) lbmol/h												
Ar	-	-	-	-	-	-	-	413	362	362	362	604
CH4	-	-	-	-	-	-	-	3	2	2	2	4
CO	6	6	6	6	-	-	-	29,041	2,820	2,820	2,820	19,827
CO_2	3	3	3	-	-	-	-	3,024	25,311	25,311	25,311	27,082
H2S	1	1	1	-	-	-	-	150	143	143	143	231
COS	-	-	-	-	-	-	-	14	-	-	-	8
HCN	-	-	-	-	-	-	-	1	-	-	-	1
H2	3	3	3	3	-	-	-	10,811	32,127	32,127	32,127	38,457
H2O	40,422	6,123	6,123	1	6,117	6,117	34,299	47,300	18,818	18,818	18,818	46,516
N2	-	-	-	-	-	-	-	456	400	400	400	667
NH3	1	-	-	-	-	-	-	9	9	9	9	14
O2	-	-	-	-	-	-	-	-	-	-	-	-
SO2	-	-	-	-	-	-	-	-	-	-	-	-
Mass Flow (Sol) lb/hr												
Coal												
Ash	5,784											
Slag												
Sulfur												
Mole Flow Rate (lbmol/hr)	40,436	6,136	6,136	10	6,117	6,117	34,299	91,222	79,993	79,993	79,993	133,411
Mass Flow Rate (lb./hr)	728,565	110,638	110,638	180	110,202	110,202	617,916	1,855,906	1,627,457	1,627,457	1,627,457	2,714,237
Temperature (F)	415.31	281.93	130.00	130.00	130.00	130.96	181,93	415.31	536.00	489.29	409.06	400.32
Pressure (PSIA)	605.34	50.76	50.76	50.76	50.76	550.00	50.76	605.34	562.69	557.69	552.69	552.69

Table A5-10. Stream mass balance for the Sour PSA coal-to-methanol case (cont.).

stream number	405	406	407	408	409	410	411	412	413	414	415	416
Description	Shifted syngas	Shifted syngas	Shifted syngas	Shifted syngas	Shifted syngas	KO water	Sour water	Sour flash gas	Sour stripper feed	Sour stripper gas	Sour gas	Sour stripper bottoms
Mole Flow (Vap/Liq) lbmol/h												
Ar	604	604	604	604	604	-	-	-	-	-	-	-
CH4	4	4	4	4	4	-	-	-	-	-	-	-
CO	19,827	19,826	19,826	19,825	19,823	2	1	-	1	1	1	-
CO_2	27,082	27,072	27,072	27,053	27,007	29	46	-	46	46	46	-
H2S	231	230	230	230	229	1	1	-	1	1	1	-
COS	8	8	8	8	8	-	-	-	-	-	-	-
HCN	1	1	1	1	-	-	1	-	-	-	-	1
H2	38,457	38,455	38,455	38,451	38,448	6	3	-	3	3	3	-
H2O	46,516	32,837	32,837	15,631	203	30,885	15,428	-	15,428	140	140	15,288
N2	667	667	667	666	666	1	0	-	-	-	-	-
NH3	14	13	13	10	-	5	10	-	-	-	-	-
O2	-	-	-	-	-	-	-	-	-	-	-	-
SO2	-	-	-	-	-	-	-	-	-	-	-	-
Mass Flow (Sol) lb/hr												
Coal												
Ash												
Slag												
Sulfur												
Mole Flow Rate (lbmol/hr)	133,411	119,716	119,716	102,483	86,993	30,928	15,490	-	15,480	192	192	15,288
Mass Flow Rate (lb./hr)	2,714,237	2,467,310	2,467,310	2,156,368	1,876,147	557,870	280,221	-	280,041	4,626	4,626	275,429
Temperature (F)	366.96	350.00	305.78	110.63	110.63	327.11	17,895.00	178.95	178.95	276.97	276.97	302.02
Pressure (PSIA)	549.69	543.24	540.34	515.83	515.83	540.34	524.53	524.53	524.53	61.11	61.11	74.16

Project: DE-FE0013363

Table A5-10. Stream mass balance for the Sour PSA coal-to-methanol case (cont.).

stream number	418	419	500	501	502	503	504	505	506	601	602
Description	Cold condesate	Hot condensate to scrubber	Cold weet syngas	Hot sweet syngas	LP CO2	HP CO2	PSA tailgas	Acid gas to claus	CO2 vent	Sulfur product	TG to coal drying
Mole Flow (Vap/Liq) lbmol/h											
Ar	-	-	583	583	22	19	22	-	3	-	3
CH4	-	-	3	3	0	0	0	-	0	-	-
CO	2	2	18,501	18,501	1,322	1,158	1,322	-	164	-	42
CO_2	30	35	494	494	26,513	23,229	26,513	598	3,284	-	595
H2S	1	1	0	0	5	4	229	228	1	-	3
COS	-	-	-	-	-	-	8	8	-	-	1
HCN	-	-	-	-	-	-	-	-	-	-	-
H2	6	6	37,183	37,183	1,265	1,108	1,265	-	157	-	15
H2O	51,885	51,890	-	-	1,857	-	203	-	-	-	375
N2	-	-	643	643	24	21	24	-	3	-	3
NH3	4	10	-	-	-	-	-	-	-	-	-
O2	-	-	-	-	-	-	-	-	-	-	-
SO2	-	-	-	-	-	-	-	-	-	-	1
Mass Flow (Sol) lb/hr											
Coal											
Ash											
Slag											
Sulfur										11,576	
Mole Flow Rate (lbmol/hr)	51,928	51,945	57,407	57,407	31,008	25,540	29,586	834	3,611	45	1,037
Mass Flow Rate (lb./hr)	936,200	936,658	656,238	656,238	1,241,580	1,058,476	1,219,909	34,581	149,656	11,576	34,569
Temperature (F)	299.48	392.00	103.54	125.00	111.00	196.00	99.01	86.00	269.76	387.64	320.00
Pressure (PSIA)	768.89	761.64	507.75	505.75	24.98	2,214.70	24.98	29.01	768.89	20.01	20.01

Project: DE-FE0013363

Table A5-10. Stream mass balance for the Sour PSA coal-to-methanol case (cont.).

stream number	701	702	703	704	705	707	708	709	710	711
Description	HP sweet syngas	cold mixed feed	Warm mixed feed	Hot mixed feed	Stage 1 product	Stage 2 cooled feed	Hot stage 2 product	Warm stage 2 product	Cold stage 2 product	Flash gas
Mole Flow (Vap/Liq) lbmol/h										
Ar	583	16,349	16,349	16,349	13,164	13,164	16,349	16,349	16,349	16,254
N2	643	19,162	19,162	19,162	15,098	15,098	19,162	19,162	19,162	19,092
CO	18,501	27,105	27,105	27,105	20,021	20,021	8,918	8,918	8,918	8,870
CO2	494	6,325	6,325	6,325	23,083	23,083	6,295	6,295	6,295	6,011
H2	37,183	58,680	58,680	58,680	38,616	38,616	22,216	22,216	22,216	22,162
H2O	0	1	1	1	500	500	39	39	39	1
CH4	3	86	86	86	77	77	86	86	86	85
CH3OH	-	1,430	1,430	1,430	14,687	14,687	19,631	19,631	19,631	1,474
C2H6	-	18	18	18	26	26	26	26	26	19
C3H8	-	-	-	-	1	1	-	-	-	-
Mole Flow Rate (lbmol/hr)	57,407	129,155	129,155	129,155	125,274	125,274	92,721	92,721	92,721	73,968
Mass Flow Rate (lb./hr)	656,238	2,393,352	2,393,352	2,393,352	3,084,930	3,084,930	2,393,233	2,393,233	2,393,233	1,790,843
Temperature (F)	228.60	177.53	437.73	400.00	475.00	400.00	430.00	230.00	130.00	130.00
Pressure (PSIA)	755.00	755.00	750.00	747.00	737.00	732.00	727.00	722.00	720.00	717.00

Table A5-10. Stream mass balance for the Sour PSA coal-to-methanol case (cont.).

stream number	712	713	714	715	716	718	724	726	727	728	730
Description	LP recycle gas	HP recycle gas	Raw methanol	PSA feed	Raw methanol	PSA product	Methanol flash gas	purge gas	Purge gas to boiler	Purge gas to coal drying	Stack
Mole Flow (Vap/Liq) lbmol/h											
Ar	15,766	15,766	95	488	4	-	91	488	338	150	589
N2	18,519	18,519	70	573	2	-	68	573	396	176	2,505
CO	8,604	8,604	48	266	2	-	46	266	184	82	-
CO2	5,831	5,831	283	180	85	-	199	180	125	56	343
H2	21,497	21,497	54	665	1	432	53	233	161	72	-
H2O	1	1	38	0	38	-	0	0	0	0	256
CH4	82	82	1	3	0	-	1	3	2	1	-
CH3OH	1,430	1,430	18,157	44	18,054	-	104	44	31	14	-
C2H6	18	18	7	1	6	-	1	1	0	0	-
C3H8	-	-	-	-	-	-	-	-	-	-	-
Mole Flow Rate (lbmol/hr)	71,749	71,749	18,753	2,219	18,190	432	563	1,787	1,237	550	3,693
Mass Flow Rate (lb./hr)	1,737,118	1,737,114	602,390	53,725	583,325	871	19,065	52,854	36,587	16,267	113,395
Temperature (F)	130.00	138.50	130.00	130.00	117.46	70.00	117.46	70.00	70.00	70.00	246.00

Project: DE-FE0013363

Table A5-11. Total plant cost summary for the coal-to-methanol reference case.

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Eng. CM H.O. & Fee	Process Contingencies %	Process Contingencies $ x1000	Project Contingencies %	Project Contingencies $ x1000	TOTAL PLANT COST $ x1000	TOTAL PLANT COST $/Gal/h
1	COAL & SORBENT HANDLING	$27,794	$4,984	$21,545	$0	$0	$54,323	$4,931	0%	$0	20%	$11,851	$71,105	$1
1.1	Coal Receive & Unload	$7,299	$0	$3,564	$0	$0	$10,864	$973	0%	$0	20%	$2,367	$14,204	$0
1.2	Coal Stackout & Reclaim	$9,432	$0	$2,286	$0	$0	$11,718	$1,027	0%	$0	20%	$2,549	$15,294	$0
1.3	Coal Conveyors & Yd Crush	$8,769	$0	$2,262	$0	$0	$11,031	$968	0%	$0	20%	$2,400	$14,399	$0
1.4	Other Coal Handling	$2,294	$0	$524	$0	$0	$2,818	$246	0%	$0	20%	$613	$3,677	$0
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.9	Coal & Sorbent Hnd.Foundations	$0	$4,984	$12,909	$0	$0	$17,892	$1,716	0%	$0	20%	$3,922	$23,530	$0
2	COAL & SORBENT PREP AND FEED	$329,508	$23,745	$52,635	$0	$0	$405,887	$35,195	0%	$0	20%	$88,216	$529,299	$6
2.1	Coal Crushing & Drying	$132,656	$6,892	$18,097	$0	$0	$157,644	$13,603	0%	$0	20%	$34,249	$205,496	$2
2.2	Prepared Coal Storage & Feed	$5,704	$1,179	$838	$0	$0	$7,721	$661	0%	$0	20%	$1,676	$10,058	$0
2.3	Dry Coal Injection System	$187,692	$1,885	$16,319	$0	$0	$205,896	$17,734	0%	$0	20%	$44,726	$268,356	$3
2.4	Misc.Coal Prep & Feed	$3,456	$2,175	$7,059	$0	$0	$12,690	$1,166	0%	$0	20%	$2,771	$16,627	$0
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$11,614	$10,323	$0	$0	$21,937	$2,031	0%	$0	20%	$4,794	$28,761	$0
3	FEEDWATER & MISC. BOP SYSTEMS	$23,901	$6,968	$15,188	$0	$0	$46,057	$4,363	0%	$0	24%	$11,966	$62,386	$1
3.1	Feedwater System	$5,414	$3,792	$2,965	$0	$0	$12,171	$1,128	0%	$0	20%	$2,660	$15,958	$0
3.2	Water Makeup & Pretreating	$1,922	$81	$648	$0	$0	$2,651	$253	0%	$0	30%	$871	$3,775	$0
3.3	Other Feedwater Subsystems	$2,962	$408	$545	$0	$0	$3,914	$352	0%	$0	20%	$853	$5,120	$0
3.4	Service Water Systems	$1,100	$922	$4,741	$0	$0	$6,763	$659	0%	$0	30%	$2,227	$9,648	$0
3.5	Other Boiler Plant Systems	$5,976	$945	$3,468	$0	$0	$10,389	$984	0%	$0	20%	$2,275	$13,647	$0
3.6	FO Supply Sys & Nat Gas	$856	$658	$909	$0	$0	$2,423	$234	0%	$0	20%	$532	$3,189	$0
3.7	Waste Treatment Equipment	$2,685	$0	$988	$0	$0	$3,673	$359	0%	$0	30%	$1,210	$5,242	$0
3.8	Misc. Power Plant Equipment	$2,987	$162	$925	$0	$0	$4,074	$393	0%	$0	30%	$1,340	$5,807	$0
4	GASIFIER & ACCESSORIES	$809,713	$26,344	$180,421	$0	$0	$1,016,478	$94,773	7%	$71,209	14%	$159,891	$1,342,351	$15
4.1	Gasifier, Syngas Cooler & Auxiliaries (Siemens)	$320,909	$0	$153,815	$0	$0	$474,724	$42,409	15%	$71,209	15%	$88,251	$676,594	$8
4.2	Syngas Cooling	w/4.1	$0	w/4.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.3	ASU/Oxidant Compression	$446,986	$0	w/equip	$0	$0	$446,986	$43,326	0%	$0	10%	$49,031	$539,342	$6
4.4	LT Heat Recovery & FG Saturation	$41,818	$0	$14,638	$0	$0	$56,457	$5,510	0%	$0	20%	$12,393	$74,360	$1
4.5	Misc. Gasification Equipment	w/4.1 &4.2	$0	w/4.1&4.2	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.6	Flare Stack System	$0	$3,494	$962	$0	$0	$4,456	$428	0%	$0	20%	$977	$5,860	$0
4.8	Major Component Rigging	w/4.1 &4.2	$0	w/4.1&4.2	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.9	Gasification Foundations	$0	$22,850	$11,006	$0	$0	$33,855	$3,100	0%	$0	25%	$9,239	$46,194	$1
5A	GAS CLEANUP & PIPING	$303,836	$6,729	$49,908	$0	$0	$360,474	$34,827	17%	$61,389	20%	$91,632	$548,323	$6
5A.1	Rectisol System	$262,792	$0	$43,278	$0	$0	$306,071	$29,601	20%	$61,214	20%	$79,377	$476,263	$5
5A.2	Elemental Sulfur Plant	$16,773	$2,469	$4,200	$0	$0	$23,442	$2,277	0%	$0	20%	$5,144	$30,863	$0
5A.3	Mercury Removal	$3,052	$0	$451	$0	$0	$3,503	$338	5%	$175	20%	$803	$4,819	$0
5A.4	Shift Reactors	$16,320	$0	$1,275	$0	$0	$17,595	$1,687	0%	$0	20%	$3,856	$23,138	$0
5A.5	Particulate Removal	w4.1	$0	w4.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
5A.6	Blowback Gas Systems	$4,899	$609	$90	$0	$0	$5,598	$531	0%	$0	20%	$1,226	$7,355	$0
5A.7	Fuel Gas Piping	$0	$1,327	$244	$0	$0	$1,571	$146	0%	$0	20%	$343	$2,060	$0
5A.9	HGCU Foundations	$0	$2,325	$369	$0	$0	$2,694	$247	0%	$0	30%	$883	$3,824	$0
5B	CO2 COMPRESSION	$97,883	$0	$34,012	$0	$0	$131,895	$12,698	0%	$0	20%	$28,919	$173,512	$2
5B.1	CO2 Removal System	w/5A.1	$0	w/5A.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
5B.2	CO2 Compression & Drying	$97,883	$0	$34,012	$0	$0	$131,895	$12,698	0%	$0	20%	$28,919	$173,512	$2
5C	MeOH Plant	$97,851	$41,167	$82,336	$0	$0	$221,354	$22,136	0%	$0	20%	$48,698	$292,188	$3
6	COMBUSTION TURBINE/ACCESSORIES	$29,881	$270	$2,244	$0	$0	$32,395	$3,028	0%	$0	10%	$3,665	$39,088	$0
6.1	Combustion Turbine Generator	$29,881	$0	$1,952	$0	$0	$31,833	$2,975	0%	$0	10%	$3,481	$38,289	$0
6.2	Open	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$270	$292	$0	$0	$562	$53	0%	$0	30%	$184	$799	$0
7	HRSG, DUCTING & STACK	$38,596	$1,301	$8,323	$0	$0	$48,220	$4,570	0%	$0	11%	$5,772	$58,562	$1
7.1	Heat Recovery Steam Generator	$34,717	$0	$4,778	$0	$0	$39,496	$3,755	0%	$0	10%	$4,325	$47,576	$1
7.2	Open	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
7.3	Ductwork	$0	$935	$1,408	$0	$0	$2,344	$206	0%	$0	20%	$510	$3,060	$0
7.4	Stack	$3,879	$0	$1,411	$0	$0	$5,289	$507	0%	$0	10%	$580	$6,376	$0
7.9	HRSG,Duct & Stack Foundations	$0	$366	$725	$0	$0	$1,091	$102	0%	$0	30%	$358	$1,550	$0
8	STEAM TURBINE GENERATOR	$44,946	$892	$11,597	$0	$0	$57,427	$5,534	0%	$0	15%	$9,198	$72,159	$1
8.1	Steam TG & Accessories	$28,660	$0	$3,849	$0	$0	$32,509	$3,119	0%	$0	10%	$3,563	$39,191	$0
8.2	Turbine Plant Auxiliaries	$88	$0	$294	$0	$0	$383	$37	0%	$0	10%	$42	$462	$0
8.3a	Condenser & Auxiliaries	$1,394	$0	$647	$0	$0	$2,041	$195	0%	$0	10%	$224	$2,460	$0
8.3b	Air Cooled Condenser	$12,778	$0	$3,719	$0	$0	$16,497	$1,650	0%	$0	20%	$3,629	$21,776	$0
8.4	Steam Piping	$1,992	$0	$2,035	$0	$0	$4,028	$346	0%	$0	25%	$1,094	$5,468	$0
8.9	TG Foundations	$0	$892	$1,077	$0	$0	$1,969	$187	0%	$0	30%	$647	$2,803	$0
9	COOLING WATER SYSTEM	$14,153	$15,226	$11,452	$0	$0	$40,831	$3,795	0%	$0	21%	$9,161	$53,787	$1
9.1	Cooling Towers	$9,562	$0	$1,636	$0	$0	$11,198	$1,067	0%	$0	15%	$1,840	$14,104	$0
9.2	Circulating Water Pumps	$2,475	$0	$143	$0	$0	$2,618	$221	0%	$0	15%	$426	$3,265	$0
9.3	Circ.Water System Auxiliaries	$218	$0	$29	$0	$0	$247	$23	0%	$0	15%	$40	$310	$0
9.4	Circ.Water Piping	$0	$10,090	$2,219	$0	$0	$12,309	$1,113	0%	$0	20%	$2,685	$16,107	$0
9.5	Make-up Water System	$822	$0	$1,105	$0	$0	$1,926	$186	0%	$0	20%	$423	$2,535	$0
9.6	Component Cooling Water Sys	$1,076	$1,425	$859	$0	$0	$3,361	$315	0%	$0	20%	$735	$4,410	$0
9.9	Circ.Water System Foundations	$0	$3,711	$5,460	$0	$0	$9,171	$870	0%	$0	30%	$3,013	$13,054	$0
10	ASH/SPENT SORBENT HANDLING SYS	$61,061	$2,818	$57,580	$0	$0	$121,459	$11,660	0%	$0	11%	$14,354	$147,472	$2
10.1	Slag Dewatering & Cooling	$53,620	$0	$50,246	$0	$0	$103,865	$9,981	0%	$0	10%	$11,385	$125,230	$1
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.4	High Temperature Ash Piping	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.5	Other Ash Recovery Equipment	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.6	Ash Storage Silos	$1,687	$0	$3,489	$0	$0	$5,176	$501	0%	$0	15%	$852	$6,529	$0
10.7	Ash Transport & Feed Equipment	$2,261	$0	$1,034	$0	$0	$3,295	$309	0%	$0	15%	$541	$4,145	$0
10.8	Misc. Ash Handling Equipment	$3,493	$2,723	$2,432	$0	$0	$8,648	$823	0%	$0	15%	$1,421	$10,892	$0
10.9	Ash/Spent Sorbent Foundation	$0	$95	$379	$0	$0	$474	$46	0%	$0	30%	$156	$676	$0
11	ACCESSORY ELECTRIC PLANT	$33,722	$20,072	$34,901	$0	$0	$88,695	$7,650	0%	$0	20%	$19,046	$115,391	$1
11.1	Generator Equipment	$966	$0	$1,270	$0	$0	$2,236	$214	0%	$0	10%	$245	$2,695	$0
11.2	Station Service Equipment	$5,055	$0	$605	$0	$0	$5,661	$522	0%	$0	10%	$618	$6,801	$0
11.3	Switchgear & Motor Control	$9,345	$0	$2,261	$0	$0	$11,606	$1,076	0%	$0	15%	$1,902	$14,584	$0
11.4	Conduit & Cable Tray	$0	$6,455	$19,047	$0	$0	$25,502	$2,466	0%	$0	25%	$6,992	$34,961	$0
11.5	Wire & Cable	$0	$12,332	$7,248	$0	$0	$19,580	$1,422	0%	$0	25%	$5,250	$26,252	$0
11.6	Protective Equipment	$0	$1,050	$3,416	$0	$0	$4,466	$436	0%	$0	15%	$735	$5,637	$0
11.7	Standby Equipment	$241	$0	$312	$0	$0	$553	$53	0%	$0	15%	$91	$697	$0
11.8	Main Power Transformers	$18,115	$0	$193	$0	$0	$18,308	$1,385	0%	$0	15%	$2,954	$22,646	$0
11.9	Electrical Foundations	$0	$235	$549	$0	$0	$784	$76	0%	$0	30%	$258	$1,118	$0
12	INSTRUMENTATION & CONTROL	$12,025	$3,635	$10,085	$0	$0	$25,745	$2,321	5%	$1,287	17%	$5,136	$34,489	$0
12.1	IGCC Control Equipment	w/4.1	$0	w/4.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.2	Combustion Turbine Control	w/6.1	$0	w/6.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.4	Other Major Component Control	$1,187	$0	$1,031	$0	$0	$2,219	$210	5%	$111	15%	$381	$2,920	$0
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.6	Control Boards,Panels & Racks	$272	$0	$228	$0	$0	$500	$48	5%	$25	20%	$115	$688	$0
12.7	Computer & Accessories	$6,333	$0	$264	$0	$0	$6,596	$605	5%	$330	10%	$753	$8,285	$0
12.8	Instrument Wiring & Tubing	$0	$3,635	$5,887	$0	$0	$9,522	$808	5%	$476	25%	$2,701	$13,507	$0
12.9	Other I & C Equipment	$4,233	$0	$2,675	$0	$0	$6,908	$650	5%	$345	15%	$1,186	$9,089	$0
13	IMPROVEMENTS TO SITE	$5,868	$2,775	$15,170	$0	$0	$23,814	$2,352	0%	$0	30%	$7,850	$34,015	$0
13.1	Site Preparation	$0	$147	$4,124	$0	$0	$4,271	$425	0%	$0	30%	$1,409	$6,105	$0
13.2	Site Improvements	$0	$2,628	$4,559	$0	$0	$7,186	$709	0%	$0	30%	$2,369	$10,264	$0
13.3	Site Facilities	$5,868	$0	$6,487	$0	$0	$12,356	$1,218	0%	$0	30%	$4,072	$17,646	$0
14	BUILDINGS & STRUCTURES	$0	$7,624	$10,565	$0	$0	$18,190	$1,656	0%	$0	16%	$3,243	$23,089	$0
14.1	Combustion Turbine Area	$0	$293	$201	$0	$0	$494	$43	0%	$0	20%	$107	$644	$0
14.2	Steam Turbine Building	$0	$2,853	$4,915	$0	$0	$7,767	$715	0%	$0	15%	$1,272	$9,754	$0
14.3	Administration Building	$0	$1,038	$911	$0	$0	$1,949	$174	0%	$0	15%	$318	$2,441	$0
14.4	Circulation Water Pumphouse	$0	$177	$114	$0	$0	$291	$25	0%	$0	15%	$47	$363	$0
14.5	Water Treatment Buildings	$0	$453	$535	$0	$0	$988	$89	0%	$0	15%	$162	$1,239	$0
14.6	Machine Shop	$0	$493	$409	$0	$0	$902	$81	0%	$0	15%	$147	$1,130	$0
14.7	Warehouse	$0	$797	$622	$0	$0	$1,419	$126	0%	$0	15%	$232	$1,776	$0
14.8	Other Buildings & Structures	$0	$475	$448	$0	$0	$924	$82	0%	$0	20%	$201	$1,207	$0
14.9	Waste Treating Building & Str.	$0	$1,044	$2,412	$0	$0	$3,456	$322	0%	$0	20%	$756	$4,534	$0
	TOTAL COST	$1,930,740	$164,550	$597,962	$0	$0	$2,693,244	$251,489	5%	$133,885	17%	$518,598	$3,597,216	$37

Table A5-12. Total plant cost summary for the coal-to-methanol Sour PSA case.

Acct. No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost	Eng. CM H.O. & Fee	Process Contingencies %	Process Contingencies $ x1000	Project Contingencies %	Project Contingencies $ x1000	TOTAL PLANT COST $ x1000	TOTAL PLANT COST $/Gal/h
1	**COAL & SORBENT HANDLING**	**$27,794**	**$4,984**	**$21,545**	**$0**	**$0**	**$54,323**	**$4,931**	**0%**	**$0**	**20%**	**$11,851**	**$71,105**	**$1**
1.1	Coal Receive & Unload	$7,299	$0	$3,564	$0	$0	$10,864	$973	0%	$0	20%	$2,367	$14,204	$0
1.2	Coal Stackout & Reclaim	$9,432	$0	$2,286	$0	$0	$11,718	$1,027	0%	$0	20%	$2,549	$15,294	$0
1.3	Coal Conveyors & Yd Crush	$8,769	$0	$2,262	$0	$0	$11,031	$968	0%	$0	20%	$2,400	$14,399	$0
1.4	Other Coal Handling	$2,294	$0	$524	$0	$0	$2,818	$246	0%	$0	20%	$613	$3,677	$0
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
1.9	Coal & Sorbent Hnd.Foundations	$0	$4,984	$12,909	$0	$0	$17,892	$1,716	0%	$0	20%	$3,922	$23,530	$0
2	**COAL & SORBENT PREP AND FEED**	**$329,508**	**$23,745**	**$52,635**	**$0**	**$0**	**$405,887**	**$35,195**	**0%**	**$0**	**20%**	**$88,216**	**$529,299**	**$6**
2.1	Coal Crushing & Drying	$132,656	$6,892	$18,097	$0	$0	$157,644	$13,603	0%	$0	20%	$34,249	$205,496	$2
2.2	Prepared Coal Storage & Feed	$5,704	$1,179	$838	$0	$0	$7,721	$661	0%	$0	20%	$1,676	$10,058	$0
2.3	Dry Coal Injection System	$187,692	$1,885	$16,319	$0	$0	$205,896	$17,734	0%	$0	20%	$44,726	$268,356	$3
2.4	Misc.Coal Prep & Feed	$3,456	$2,175	$7,059	$0	$0	$12,690	$1,166	0%	$0	20%	$2,771	$16,627	$0
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$11,614	$10,323	$0	$0	$21,937	$2,031	0%	$0	20%	$4,794	$28,761	$0
3	**FEEDWATER & MISC. BOP SYSTEMS**	**$23,901**	**$6,968**	**$15,188**	**$0**	**$0**	**$46,057**	**$4,363**	**0%**	**$0**	**24%**	**$11,966**	**$62,386**	**$1**
3.1	Feedwater System	$5,414	$3,792	$2,965	$0	$0	$12,171	$1,128	0%	$0	20%	$2,660	$15,958	$0
3.2	Water Makeup & Pretreating	$1,922	$81	$648	$0	$0	$2,651	$253	0%	$0	30%	$871	$3,775	$0
3.3	Other Feedwater Subsystems	$2,962	$408	$545	$0	$0	$3,914	$352	0%	$0	20%	$853	$5,120	$0
3.4	Service Water Systems	$1,100	$922	$4,741	$0	$0	$6,763	$659	0%	$0	30%	$2,227	$9,648	$0
3.5	Other Boiler Plant Systems	$5,976	$945	$3,468	$0	$0	$10,389	$984	0%	$0	20%	$2,275	$13,647	$0
3.6	FO Supply Sys & Nat Gas	$856	$658	$909	$0	$0	$2,423	$234	0%	$0	20%	$532	$3,189	$0
3.7	Waste Treatment Equipment	$2,685	$0	$988	$0	$0	$3,673	$359	0%	$0	30%	$1,210	$5,242	$0
3.8	Misc. Power Plant Equipment	$2,987	$162	$925	$0	$0	$4,074	$393	0%	$0	30%	$1,340	$5,807	$0
4	**GASIFIER & ACCESSORIES**	**$809,713**	**$26,344**	**$180,421**	**$0**	**$0**	**$1,016,478**	**$94,773**	**7%**	**$71,209**	**14%**	**$159,891**	**$1,342,351**	**$15**
4.1	Gasifier, Syngas Cooler & Auxiliaries (Siemens)	$320,909	$0	$153,815	$0	$0	$474,724	$42,409	15%	$71,209	15%	$88,251	$676,594	$8
4.2	Syngas Cooling	w/4.1	$0	w/4.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.3	ASU/Oxidant Compression	$446,986	$0	w/equip	$0	$0	$446,986	$43,326	0%	$0	10%	$49,031	$539,342	$6
4.4	LT Heat Recovery & FG Saturation	$41,818	$0	$14,638	$0	$0	$56,457	$5,510	0%	$0	20%	$12,393	$74,360	$1
4.5	Misc. Gasification Equipment	w/4.1 &4.2	$0	w/4.1&4.2	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.6	Flare Stack System	$0	$3,494	$962	$0	$0	$4,456	$428	0%	$0	20%	$977	$5,860	$0
4.8	Major Component Rigging	w/4.1 &4.2	$0	w/4.1&4.2	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
4.9	Gasification Foundations	$0	$22,850	$11,006	$0	$0	$33,855	$3,100	0%	$0	25%	$9,239	$46,194	$1
5A	**GAS CLEANUP & PIPING**	**$107,366**	**$5,164**	**$56,641**	**$0**	**$0**	**$169,171**	**$16,340**	**13%**	**$21,628**	**19%**	**$39,159**	**$246,298**	**$3**
5A.1	Sour PSA	$52,670	$0	$44,690	$0	$0	$97,360	$9,416	20%	$19,472	20%	$25,250	$151,498	$2
5A.2	Elemental Sulfur Plant	$16,442	$2,420	$4,117	$0	$0	$22,979	$2,232	0%	$0	20%	$5,042	$30,254	$0
5A.3	Mercury Removal	$3,052	$0	$451	$0	$0	$3,503	$338	5%	$175	20%	$803	$4,819	$0
5A.4	Shift Reactors	$16,153	$0	$1,262	$0	$0	$17,415	$1,670	0%	$0	20%	$3,817	$22,902	$0
5A.5	Particulate Removal	w4.1	$0	w4.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
5A.6	Blowback Gas Systems	$4,899	$609	$90	$0	$0	$5,598	$531	0%	$0	20%	$1,226	$7,355	$0
5A.7	Fuel Gas Piping	$0	$1,327	$244	$0	$0	$1,571	$146	0%	$0	20%	$343	$2,060	$0
5A.8	AGE	$14,150	$0	$5,660	$0	$0	$19,810	$1,922	10%	$1,981	10%	$2,371	$26,084	$0
5A.9	HGCU Foundations	$0	$808	$126	$0	$0	$934	$86	0%	$0	30%	$306	$1,326	$0
5B	**CO2 COMPRESSION**	**$109,594**	**$0**	**$39,204**	**$0**	**$0**	**$148,798**	**$14,637**	**0%**	**$0**	**20%**	**$32,687**	**$196,121**	**$2**
5B.1	CO2 Removal System	w/5A.1	$0	w/5A.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
5B.2	CO2 Compression & Drying	$109,594	$0	$39,204	$0	$0	$148,798	$14,637	0%	$0	20%	$32,687	$196,121	$2
5C	MeOH Plant	**$92,986**	**$39,121**	**$78,242**	**$0**	**$0**	**$210,350**	**$21,035**	**0%**	**$0**	**20%**	**$46,277**	**$277,662**	**$3**
6	**COMBUSTION TURBINE/ACCESSORIES**	**$29,881**	**$270**	**$2,244**	**$0**	**$0**	**$32,395**	**$3,028**	**0%**	**$0**	**10%**	**$3,665**	**$39,088**	**$0**
6.1	Combustion Turbine Generator	$29,881	$0	$1,952	$0	$0	$31,833	$2,975	0%	$0	10%	$3,481	$38,289	$0
6.2	Open	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$270	$292	$0	$0	$562	$53	0%	$0	30%	$184	$799	$0
7	**HRSG, DUCTING & STACK**	**$34,318**	**$1,301**	**$7,734**	**$0**	**$0**	**$43,352**	**$4,107**	**0%**	**$0**	**11%**	**$5,239**	**$52,699**	**$1**
7.1	Heat Recovery Steam Generator	$30,439	$0	$4,189	$0	$0	$34,628	$3,292	0%	$0	10%	$3,792	$41,712	$0
7.2	Open	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
7.3	Ductwork	$0	$935	$1,408	$0	$0	$2,344	$206	0%	$0	20%	$510	$3,060	$0
7.4	Stack	$3,879	$0	$1,411	$0	$0	$5,289	$507	0%	$0	10%	$580	$6,376	$0
7.9	HRSG,Duct & Stack Foundations	$0	$366	$725	$0	$0	$1,091	$102	0%	$0	30%	$358	$1,550	$0
8	**STEAM TURBINE GENERATOR**	**$39,623**	**$780**	**$10,436**	**$0**	**$0**	**$50,839**	**$4,894**	**0%**	**$0**	**15%**	**$8,197**	**$63,930**	**$1**
8.1	Steam TG & Accessories	$25,128	$0	$3,375	$0	$0	$28,503	$2,734	0%	$0	10%	$3,124	$34,361	$0
8.2	Turbine Plant Auxiliaries	$77	$0	$257	$0	$0	$335	$33	0%	$0	10%	$37	$404	$0
8.3a	Condenser & Auxiliaries	$1,223	$0	$567	$0	$0	$1,790	$171	0%	$0	10%	$196	$2,157	$0
8.3b	Air Cooled Condenser	$11,203	$0	$3,261	$0	$0	$14,464	$1,447	0%	$0	20%	$3,182	$19,092	$0
8.4	Steam Piping	$1,992	$0	$2,035	$0	$0	$4,028	$346	0%	$0	25%	$1,094	$5,468	$0
8.9	TG Foundations	$0	$780	$941	$0	$0	$1,720	$163	0%	$0	30%	$565	$2,448	$0
9	**COOLING WATER SYSTEM**	**$14,153**	**$15,226**	**$11,452**	**$0**	**$0**	**$40,831**	**$3,795**	**0%**	**$0**	**21%**	**$9,161**	**$53,787**	**$1**
9.1	Cooling Towers	$9,562	$0	$1,636	$0	$0	$11,198	$1,067	0%	$0	15%	$1,840	$14,104	$0
9.2	Circulating Water Pumps	$2,475	$0	$143	$0	$0	$2,618	$221	0%	$0	15%	$426	$3,265	$0
9.3	Circ.Water System Auxiliaries	$218	$0	$29	$0	$0	$247	$23	0%	$0	15%	$40	$310	$0
9.4	Circ.Water Piping	$0	$10,090	$2,219	$0	$0	$12,309	$1,113	0%	$0	20%	$2,685	$16,107	$0
9.5	Make-up Water System	$822	$0	$1,105	$0	$0	$1,926	$186	0%	$0	20%	$423	$2,535	$0
9.6	Component Cooling Water Sys	$1,076	$1,425	$859	$0	$0	$3,361	$315	0%	$0	20%	$735	$4,410	$0
9.9	Circ.Water System Foundations	$0	$3,711	$5,460	$0	$0	$9,171	$870	0%	$0	30%	$3,013	$13,054	$0
10	**ASH/SPENT SORBENT HANDLING SYS**	**$61,061**	**$2,818**	**$57,580**	**$0**	**$0**	**$121,459**	**$11,660**	**0%**	**$0**	**11%**	**$14,354**	**$147,472**	**$2**
10.1	Slag Dewatering & Cooling	$53,620	$0	$50,246	$0	$0	$103,865	$9,981	0%	$0	10%	$11,385	$125,230	$1
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.3	Cleanup Ash Depressurization	w/10.1	w/10.1	w/10.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.4	High Temperature Ash Piping	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.5	Other Ash Recovery Equipment	$0	$0	$0	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
10.6	Ash Storage Silos	$1,687	$0	$3,489	$0	$0	$5,176	$501	0%	$0	15%	$852	$6,529	$0
10.7	Ash Transport & Feed Equipment	$2,261	$0	$1,034	$0	$0	$3,295	$309	0%	$0	15%	$541	$4,145	$0
10.8	Misc. Ash Handling Equipment	$3,493	$2,723	$2,432	$0	$0	$8,648	$823	0%	$0	15%	$1,421	$10,892	$0
10.9	Ash/Spent Sorbent Foundation	$0	$95	$379	$0	$0	$474	$46	0%	$0	30%	$156	$676	$0
11	**ACCESSORY ELECTRIC PLANT**	**$30,992**	**$19,308**	**$33,526**	**$0**	**$0**	**$83,826**	**$7,244**	**0%**	**$0**	**20%**	**$18,072**	**$109,142**	**$1**
11.1	Generator Equipment	$872	$0	$1,147	$0	$0	$2,020	$193	0%	$0	10%	$221	$2,435	$0
11.2	Station Service Equipment	$4,857	$0	$582	$0	$0	$5,439	$502	0%	$0	10%	$594	$6,534	$0
11.3	Switchgear & Motor Control	$8,978	$0	$2,172	$0	$0	$11,150	$1,034	0%	$0	15%	$1,828	$14,012	$0
11.4	Conduit & Cable Tray	$0	$6,202	$18,299	$0	$0	$24,501	$2,369	0%	$0	25%	$6,718	$33,588	$0
11.5	Wire & Cable	$0	$11,848	$6,964	$0	$0	$18,811	$1,366	0%	$0	25%	$5,044	$25,222	$0
11.6	Protective Equipment	$0	$1,050	$3,416	$0	$0	$4,466	$436	0%	$0	15%	$735	$5,637	$0
11.7	Standby Equipment	$222	$0	$288	$0	$0	$510	$49	0%	$0	15%	$84	$643	$0
11.8	Main Power Transformers	$16,062	$0	$171	$0	$0	$16,233	$1,228	0%	$0	15%	$2,619	$20,080	$0
11.9	Electrical Foundations	$0	$208	$488	$0	$0	$696	$67	0%	$0	30%	$229	$993	$0
12	**INSTRUMENTATION & CONTROL**	**$11,887**	**$3,593**	**$9,969**	**$0**	**$0**	**$25,449**	**$2,294**	**5%**	**$1,272**	**17%**	**$5,077**	**$34,092**	**$0**
12.1	IGCC Control Equipment	w/4.1	$0	w/4.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.2	Combustion Turbine Control	w/6.1	$0	w/6.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.4	Other Major Component Control	$1,174	$0	$1,019	$0	$0	$2,193	$207	5%	$110	15%	$377	$2,887	$0
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	0%	$0	0%	$0	$0	$0
12.6	Control Boards,Panels & Racks	$269	$0	$225	$0	$0	$495	$47	5%	$25	20%	$113	$680	$0
12.7	Computer & Accessories	$6,260	$0	$261	$0	$0	$6,520	$598	5%	$326	10%	$744	$8,189	$0
12.8	Instrument Wiring & Tubing	$0	$3,593	$5,819	$0	$0	$9,412	$798	5%	$471	25%	$2,670	$13,352	$0
12.9	Other I & C Equipment	$4,184	$0	$2,644	$0	$0	$6,828	$643	5%	$341	15%	$1,172	$8,985	$0
13	**IMPROVEMENTS TO SITE**	**$5,832**	**$2,758**	**$15,076**	**$0**	**$0**	**$23,666**	**$2,337**	**0%**	**$0**	**30%**	**$7,801**	**$33,804**	**$0**
13.1	Site Preparation	$0	$146	$4,098	$0	$0	$4,245	$422	0%	$0	30%	$1,400	$6,067	$0
13.2	Site Improvements	$0	$2,611	$4,530	$0	$0	$7,142	$705	0%	$0	30%	$2,354	$10,201	$0
13.3	Site Facilities	$5,832	$0	$6,447	$0	$0	$12,279	$1,210	0%	$0	30%	$4,047	$17,536	$0
14	**BUILDINGS & STRUCTURES**	**$0**	**$7,576**	**$10,491**	**$0**	**$0**	**$18,067**	**$1,645**	**0%**	**$0**	**16%**	**$3,223**	**$22,935**	**$0**
14.1	Combustion Turbine Area	$0	$293	$201	$0	$0	$494	$43	0%	$0	20%	$107	$644	$0
14.2	Steam Turbine Building	$0	$2,815	$4,850	$0	$0	$7,665	$705	0%	$0	15%	$1,256	$9,626	$0
14.3	Administration Building	$0	$1,030	$904	$0	$0	$1,934	$173	0%	$0	15%	$316	$2,423	$0
14.4	Circulation Water Pumphouse	$0	$177	$114	$0	$0	$291	$25	0%	$0	15%	$47	$363	$0
14.5	Water Treatment Buildings	$0	$453	$535	$0	$0	$988	$89	0%	$0	15%	$162	$1,239	$0
14.6	Machine Shop	$0	$493	$408	$0	$0	$901	$81	0%	$0	15%	$147	$1,128	$0
14.7	Warehouse	$0	$795	$621	$0	$0	$1,417	$125	0%	$0	15%	$231	$1,773	$0
14.8	Other Buildings & Structures	$0	$475	$447	$0	$0	$922	$82	0%	$0	20%	$201	$1,205	$0
14.9	Waste Treating Building & Str.	$0	$1,044	$2,412	$0	$0	$3,456	$322	0%	$0	20%	$756	$4,534	$0
	TOTAL COST	**$1,728,609**	**$159,954**	**$602,384**	**$0**	**$0**	**$2,490,946**	**$232,278**	**4%**	**$94,109**	**16%**	**$464,836**	**$3,282,170**	**$34**

Table A5-13. O&M cost summary for the coal-to-methanol reference case.

Case	4 shell gasifiers			
Plant size	89,459	Gal/h		
			Capacity factor	90%

Operating and Maintenance Labor

Operating Labor				
Operating Labor Rate	39.7	$/hour		
Operating Labor Burden Rate	30%	of base		
Labor Over Head Charge Rate	25%	of labor		
Administrative & Support Labor	25%	of burdened O&M labor		
Maintenance labor	0.964%	of TPC*		
Work force	**16**			
Skilled Operator	2			
Operator	10			
Foreman	1			
Lab Technician, etc	3			
			Annual Cost ($)	**Annual Unit Cost ($/bbl)**
Fixed Operating Costs				
Operating Labor			$8,137,865	$0.48461
Maintenance Labor			$34,662,770	$2.06416
Administrative & Support Labor			$10,700,159	$0.63719
Property Taxes & Insurance			$71,944,314	$4.28427
TOTAL FIXED OPERATING COST			**$125,445,107**	**$7.47022**

* (back calculated from NETL)

Variable Operating Costs

					Annual Cost ($)	**Annual Unit Cost ($/bbl)**	
Maintenance Material Cost	1.45%	of TPC*			**$ 51,994,155**	**$3.09624**	
	Consumption						
	Initial	**per day**	**Unit Cost ($)**	**Initial Cost ($)**	**Annual Cost ($)**	**Annual Unit Cost ($/bbl)**	
Water (1000 gallons)	0.00	5,149	1.67	**$0**	**$2,831,566**	**$0.16862**	
Chemicals				**$5,960,050**	**$5,043,540**	**$0.30034**	
MU & WT Chem (lb)	0	30,678	0.27	$0	2,699,144	$0.16073	
Carbon (Mercury Removal) (lb)	169,892	259	1.63	$276,116	138,057	$0.00822	
COS Catalyst (m3)	0	0	3751.70	$0	0	$0.00000	
Water Gas Shift Catalyst (ft3)	6,793	4.65	771.99	$5,244,339	1,179,976	$0.07027	
EconamineFG+ MEA (ton)	117	0.16	3751.70	$439,595	202,615	$0.01207	
MeOH Synthesis Catalyst (ft3)	0	0.00	534.67	$0	0	$0.00000	
TEG for CO2 dryer (Gal)	0	381	6.57	$0	823,747	$0.04905	
Claus Catalyst (ft3)	0	0.00	203.15	$0	0	$0.00000	
Other				**$0**	**$0**	**$0.00000**	
Supplemental Fuel (MMBtu)	0	0		$0	$0	$0.00000	
Gases, N2 etc (100 scf)	0	0		$0	$0	$0.00000	
L.P. Steam (1000 lb)	0	0		$0	$0	$0.00000	
Waste Disposal				**$0**	**$47,500,802**	**$2.82866**	
Spent Mercury Catalyst (lb)	0	259	0.65	$0	$55,223	$0.00329	
Flyash (ton)	0	0	0.00	$0	$0	$0.00000	
Slag (ton)	0	1,340	25.11	$0	$11,051,900	$0.65814	
CO2 TS&M (1000 kg)	*0*	*10,072*	*11.00*	*$0*	*$36,393,679*	$2.16723	
By-products and Emissions				**$0**	**($19,889,108)**	**($1.18439)**	
Sulfur (ton)	0	0	0	$0	$0	$0.00000	
power(MW.h)	0	1,016.03	-59.59		$ (19,889,108)	($1.18439)	
TOTAL VARIABLE OPERATING COSTS				**$5,960,050**	**$87,480,955**	**$5.20947**	$3.04223
Fuel				**$0**	**$207,435,799**	**$12.35275**	
Coal (ton)	0	12,501	36.57	$0	$150,176,134	$8.94295	
Natural Gas (MMBTU)	0	28,435	6.13	$0	$57,259,665	$3.40980	

Project: DE-FE0013363

Table A5-14. O&M cost summary for the coal-to-methanol Sour PSA case.

Operating and Maintenance Labor

			Annual Cost ($)	Annual Unit Cost ($/bbl)
Operating Labor				
Operating Labor Rate	39.7	$/hour		
Operating Labor Burden Rate	30%	of base		
Labor Over Head Charge Rate	25%	of labor		
Administrative & Support Labor	25%	of burdened O&M labor		
Maintenance labor	0.964%	of TPC*		
Work force	**16**			
Skilled Operator	2			
Operator	10			
Foreman	1			
Lab Technician, etc	3			
Fixed Operating Costs				
Operating Labor			$8,137,865	$0.49467
Maintenance Labor			$31,626,990	$1.92249
Administrative & Support Labor			$9,941,214	$0.60429
Property Taxes & Insurance			$65,643,400	$3.99022
TOTAL FIXED OPERATING COST			**$115,349,469**	**$7.01166**

* (back calculated from NETL)

Variable Operating Costs

					Annual Cost ($)	Annual Unit Cost ($/bbl)
Maintenance Material Cost	1.45%	of TPC*			**$ 47,440,485**	**$2.88373**
	Consumption Initial	**per day**	**Unit Cost ($)**	**Initial Cost ($)**	**Annual Cost ($)**	**Annual Unit Cost ($/bbl)**
Water (1000 gallons)	0.00	5,149	1.67	**$0**	**$2,831,566**	**$0.17212**
Chemicals				**$21,970,071**	**$7,388,076**	**$0.44909**
MU & WT Chem (lb)	0	30,678	0.27	$0	2,699,144	$0.16407
Carbon (Mercury Removal) (lb)	169,892	259	1.63	$276,116	138,057	$0.00839
COS Catalyst (m3)	0	0	3751.70	$0	0	$0.00000
Water Gas Shift Catalyst (ft3)	6,793	4.65	771.99	$5,244,339	1,179,976	$0.07173
EconamineFG+ MEA (ton)	117	0.16	3751.70	$439,595	202,615	$0.01232
MeOH Synthesis Catalyst (ft3)	2,428	2.22	534.67	$1,297,933	389,380	$0.02367
TEG for CO2 dryer (Gal)	0	448	6.57	$0	968,071	$0.05885
AGE (Flexorb) Solution (Gal)				$6,690,000	1,003,500	$0.06100
Sour PSA adsorbant (lb)	7,180,530	1,967	1.12	$8,022,088	721,988	$0.04389
Claus Catalyst (ft3)	0	1.28	203.15	$0	85,344	$0.00519
Other				**$0**	**$0**	**$0.00000**
Supplemental Fuel (MMBtu)	0	0		$0	$0	$0.00000
Gases, N2 etc (100 scf)	0	0		$0	$0	$0.00000
L.P. Steam (1000 lb)	0	0		$0	$0	$0.00000
Waste Disposal				**$0**	**$53,907,698**	**$3.27685**
Spent Mercury Catalyst (lb)	0	402	0.65	$0	$85,781	$0.00521
Flyash (ton)	0	0	0.00	$0	$0	$0.00000
Slag (ton)	0	1,340	25.11	$0	$11,051,900	$0.67180
CO2 TS&M (1000 kg)	*0*	*11,836*	*11.00*	*$0*	*$42,770,017*	$2.59983
By-products and Emissions				**$0**	**($14,571,577)**	**($0.88575)**
Sulfur (ton)	0	87	0	$0	$0	$0.00000
power(MW.h)	0	744.39	-59.59		$ (14,571,577)	($0.88575)
TOTAL VARIABLE OPERATING COSTS				**$21,970,071**	**$96,996,248**	**$5.89604**
Fuel				**$0**	**$207,435,799**	**$12.60924**
Coal (ton)	0	12,501	36.57	$0	$150,176,134	$9.12864
Natural Gas (MMBTU)	0	28,435	6.13	$0	$57,259,665	$3.48060

Table A5-15. Cash flow analysis for the reference coal-to-methanol case, in $ x1000.

plant net output	89,459	Gal/h
Construction period	5	year
estimated MeOH RSP	1.89	$/gal

end of	year 1	year 2	year 3	year 4	year 5	year 6	year 7	year 8	year 9	year 10	year 11	year 12	year 13	year 14	year 15
Operating revenues	$ -	$ -	$ -	$ -	$ -	$ 1,332,875	$ 1,372,862	$ 1,414,048	$ 1,456,469	$ 1,500,163	$ 1,545,168	$ 1,591,523	$ 1,639,269	$ 1,688,447	$ 1,739,100
Operating expenses															
Fixed	$ -	$ -	$ -	$ -	$ -	$ 145,425	$ 149,788	$ 154,282	$ 158,910	$ 163,677	$ 168,588	$ 173,645	$ 178,855	$ 184,220	$ 189,747
Variable	$ -	$ -	$ -	$ -	$ -	$ 101,414	$ 104,457	$ 107,591	$ 110,818	$ 114,143	$ 117,567	$ 121,094	$ 124,727	$ 128,469	$ 132,323
Fuel	$ -	$ -	$ -	$ -	$ -	$ 240,475	$ 247,689	$ 255,120	$ 262,773	$ 270,657	$ 278,776	$ 287,140	$ 295,754	$ 304,626	$ 313,765
Operating income	$ -	$ -	$ -	$ -	$ -	$ 845,561	$ 870,928	$ 897,055	$ 923,967	$ 951,686	$ 980,237	$ 1,009,644	$ 1,039,933	$ 1,071,131	$ 1,103,265
Interest Expense	$ -	$ -	$ -	$ -	$ -	$ 175,908	$ 174,356	$ 172,679	$ 170,867	$ 168,911	$ 166,799	$ 164,517	$ 162,053	$ 159,392	$ 156,517
Depreciation & Amortisation	$ -	$ -	$ -	$ -	$ -	$ 164,358	$ 316,400	$ 292,644	$ 270,730	$ 250,393	$ 231,635	$ 214,235	$ 198,193	$ 195,564	$ 195,520
Taxable Income	$ -	$ -	$ -	$ -	$ -	$ 505,295	$ 380,173	$ 431,733	$ 482,370	$ 532,381	$ 581,803	$ 630,892	$ 679,687	$ 716,176	$ 751,228
Income Taxes	$ -	$ -	$ -	$ -	$ -	$ 192,012	$ 144,466	$ 164,058	$ 183,301	$ 202,305	$ 221,085	$ 239,739	$ 258,281	$ 272,147	$ 285,467
Net Income	$ -	$ -	$ -	$ -	$ -	$ 313,283	$ 235,707	$ 267,674	$ 299,069	$ 330,077	$ 360,718	$ 391,153	$ 421,406	$ 444,029	$ 465,761
Cash form Operation	$ -	$ -	$ -	$ -	$ -	$ 845,561	$ 870,928	$ 897,055	$ 923,967	$ 951,686	$ 980,237	$ 1,009,644	$ 1,039,933	$ 1,071,131	$ 1,103,265
Income Taxes	$ -	$ -	$ -	$ -	$ -	$ 192,012	$ 144,466	$ 164,058	$ 183,301	$ 202,305	$ 221,085	$ 239,739	$ 258,281	$ 272,147	$ 285,467
Total Interest Expense	$ -	$ -	$ -	$ -	$ -	$ 175,908	$ 174,356	$ 172,679	$ 170,867	$ 168,911	$ 166,799	$ 164,517	$ 162,053	$ 159,392	$ 156,517
Total Principal Repayment	$ -	$ -	$ -	$ -	$ -	$ 19,410	$ 20,963	$ 22,640	$ 24,451	$ 26,407	$ 28,520	$ 30,802	$ 33,266	$ 35,927	$ 38,801
Operating Cash Flow	$ -	$ -	$ -	$ -	$ -	$ 458,230	$ 531,143	$ 537,678	$ 545,348	$ 554,063	$ 563,833	$ 574,586	$ 586,334	$ 603,666	$ 622,480
Capital Cost	$ 359,722	$ 1,118,015	$ 965,219	$ 799,974	$ 621,580	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ (359,722)	$ (1,118,015)	$ (965,219)	$ (799,974)	$ (621,580)	$ 458,230	$ 531,143	$ 537,678	$ 545,348	$ 554,063	$ 563,833	$ 574,586	$ 586,334	$ 603,666	$ 622,480
Loan Draws	$ 179,860.78	$ 559,007.32	$ 482,609.65	$ 399,986.88	$ 310,789.80	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ (179,861)	$ (559,007)	$ (482,610)	$ (399,987)	$ (310,790)	$ 458,230	$ 531,143	$ 537,678	$ 545,348	$ 554,063	$ 563,833	$ 574,586	$ 586,334	$ 603,666	$ 622,480
Equity Draws	$ 179,861	$ 559,007	$ 482,610	$ 399,987	$ 310,790	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ -	$ -	$ -	$ -	$ -	$ 458,230	$ 531,143	$ 537,678	$ 545,348	$ 554,063	$ 563,833	$ 574,586	$ 586,334	$ 603,666	$ 622,480

Internal rate of Return	**20.00%**
Net Present Value at discount rate	
8%	$ 3,652,241
10%	$ 2,417,528
12%	$ 1,570,022
14%	$ 974,214
20%	$ 0

Table A5-15. Cash flow analysis for the reference coal-to-methanol case, in $ x1000 (cont.).

end of	year 16	year 17	year 18	year 19	year 20	year 21	year 22	year 23	year 24	year 25
Operating revenues	$ 1,791,273	$ 1,845,011	$ 1,900,362	$ 1,957,373	$ 2,016,094	$ 2,076,577	$ 2,138,874	$ 2,203,040	$ 2,269,131	$ 2,337,205
Operating expenses										
Fixed	$ 195,439	$ 201,303	$ 207,342	$ 213,562	$ 219,969	$ 226,568	$ 233,365	$ 240,366	$ 247,577	$ 255,004
Variable	$ 136,292	$ 140,381	$ 144,593	$ 148,930	$ 153,398	$ 158,000	$ 162,740	$ 167,623	$ 172,651	$ 177,831
Fuel	$ 323,178	$ 332,874	$ 342,860	$ 353,146	$ 363,740	$ 374,652	$ 385,892	$ 397,468	$ 409,392	$ 421,674
Operating income	$ 1,136,363	$ 1,170,454	$ 1,205,568	$ 1,241,735	$ 1,278,987	$ 1,317,356	$ 1,356,877	$ 1,397,583	$ 1,439,511	$ 1,482,696
Interest Expense	$ 153,413	$ 150,061	$ 146,440	$ 142,530	$ 138,307	$ 133,746	$ 128,820	$ 123,500	$ 117,755	$ 111,550
Depreciation & Amortisation	$ 195,564	$ 195,520	$ 195,564	$ 195,520	$ 195,564	$ 195,520	$ 195,564	$ 195,520	$ 195,564	$ 195,520
Taxable Income	$ 787,386	$ 824,873	$ 863,563	$ 903,685	$ 945,116	$ 988,090	$ 1,032,493	$ 1,078,563	$ 1,126,192	$ 1,175,626
Income Taxes	$ 299,207	$ 313,452	$ 328,154	$ 343,400	$ 359,144	$ 375,474	$ 392,347	$ 409,854	$ 427,953	$ 446,738
Net Income	$ 488,179	$ 511,421	$ 535,409	$ 560,284	$ 585,972	$ 612,616	$ 640,146	$ 668,709	$ 698,239	$ 728,888
Cash form Operation	$ 1,136,363	$ 1,170,454	$ 1,205,568	$ 1,241,735	$ 1,278,987	$ 1,317,356	$ 1,356,877	$ 1,397,583	$ 1,439,511	$ 1,482,696
Income Taxes	$ 299,207	$ 313,452	$ 328,154	$ 343,400	$ 359,144	$ 375,474	$ 392,347	$ 409,854	$ 427,953	$ 446,738
Total Interest Expense	$ 153,413	$ 150,061	$ 146,440	$ 142,530	$ 138,307	$ 133,746	$ 128,820	$ 123,500	$ 117,755	$ 111,550
Total Principal Repayment	$ 41,905	$ 45,258	$ 48,878	$ 52,789	$ 57,012	$ 61,573	$ 66,498	$ 71,818	$ 77,564	$ 83,769
Operating Cash Flow	$ 641,838	$ 661,684	$ 682,095	$ 703,016	$ 724,524	$ 746,563	$ 769,211	$ 792,411	$ 816,239	$ 840,639
Capital Cost	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ 641,838	$ 661,684	$ 682,095	$ 703,016	$ 724,524	$ 746,563	$ 769,211	$ 792,411	$ 816,239	$ 840,639
Loan Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ 641,838	$ 661,684	$ 682,095	$ 703,016	$ 724,524	$ 746,563	$ 769,211	$ 792,411	$ 816,239	$ 840,639
Equity Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ 641,838	$ 661,684	$ 682,095	$ 703,016	$ 724,524	$ 746,563	$ 769,211	$ 792,411	$ 816,239	$ 840,639

end of	year 26	year 27	year 28	year 29	year 30	year 31	year 32	year 33	year 34	year 35
Operating revenues	$ 2,407,321	$ 2,479,541	$ 2,553,927	$ 2,630,545	$ 2,709,461	$ 2,790,745	$ 2,874,468	$ 2,960,702	$ 3,049,523	$ 3,141,008
Operating expenses										
Fixed	$ 262,654	$ 270,534	$ 278,650	$ 287,009	$ 295,620	$ 304,488	$ 313,623	$ 323,032	$ 332,722	$ 342,704
Variable	$ 183,166	$ 188,661	$ 194,320	$ 200,150	$ 206,155	$ 212,339	$ 218,709	$ 225,271	$ 232,029	$ 238,990
Fuel	$ 434,324	$ 447,354	$ 460,775	$ 474,598	$ 488,836	$ 503,501	$ 518,606	$ 534,164	$ 550,189	$ 566,695
Operating income	$ 1,527,177	$ 1,572,992	$ 1,620,182	$ 1,668,787	$ 1,718,851	$ 1,770,417	$ 1,823,529	$ 1,878,235	$ 1,934,582	$ 1,992,619
Interest Expense	$ 104,848	$ 97,611	$ 89,794	$ 81,352	$ 72,235	$ 62,388	$ 51,754	$ 40,268	$ 27,864	$ 14,468
Depreciation & Amortisation	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Taxable Income	$ 1,422,329	$ 1,475,382	$ 1,530,388	$ 1,587,435	$ 1,646,616	$ 1,708,029	$ 1,771,776	$ 1,837,967	$ 1,906,718	$ 1,978,151
Income Taxes	$ 540,485	$ 560,645	$ 581,547	$ 603,225	$ 625,714	$ 649,051	$ 673,275	$ 698,427	$ 724,553	$ 751,698
Net Income	$ 881,844	$ 914,737	$ 948,841	$ 984,210	$ 1,020,902	$ 1,058,978	$ 1,098,501	$ 1,139,539	$ 1,182,165	$ 1,226,454
Cash form Operation	$ 1,527,177	$ 1,572,992	$ 1,620,182	$ 1,668,787	$ 1,718,851	$ 1,770,417	$ 1,823,529	$ 1,878,235	$ 1,934,582	$ 1,992,619
Income Taxes	$ 540,485	$ 560,645	$ 581,547	$ 603,225	$ 625,714	$ 649,051	$ 673,275	$ 698,427	$ 724,553	$ 751,698
Total Interest Expense	$ 104,848	$ 97,611	$ 89,794	$ 81,352	$ 72,235	$ 62,388	$ 51,754	$ 40,268	$ 27,864	$ 14,468
Total Principal Repayment	$ 90,470	$ 97,708	$ 105,525	$ 113,967	$ 123,084	$ 132,931	$ 143,565	$ 155,050	$ 167,454	$ 180,851
Operating Cash Flow	$ 791,373	$ 817,029	$ 843,316	$ 870,243	$ 897,818	$ 926,047	$ 954,936	$ 984,489	$ 1,014,711	$ 1,045,603
Capital Cost	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ 791,373	$ 817,029	$ 843,316	$ 870,243	$ 897,818	$ 926,047	$ 954,936	$ 984,489	$ 1,014,711	$ 1,045,603
Loan Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ 791,373	$ 817,029	$ 843,316	$ 870,243	$ 897,818	$ 926,047	$ 954,936	$ 984,489	$ 1,014,711	$ 1,045,603
Equity Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ 791,373	$ 817,029	$ 843,316	$ 870,243	$ 897,818	$ 926,047	$ 954,936	$ 984,489	$ 1,014,711	$ 1,045,603

Table A5-16. Cash flow analysis for the Sour PSA coal-to-methanol case, in $ x1000.

plant net output	87,639	Gal/h
Construction period	5	year
estimated MeOH RSP	1.82	$/gal

end of	year 1	year 2	year 3	year 4	year 5	year 6	year 7	year 8	year 9	year 10	year 11	year 12	year 13	year 14	year 15
Operating revenues	$ -	$ -	$ -	$ -	$ -	$ 1,258,948	$ 1,296,717	$ 1,335,618	$ 1,375,687	$ 1,416,957	$ 1,459,466	$ 1,503,250	$ 1,548,348	$ 1,594,798	$ 1,642,642
Operating expenses															
Fixed	$ -	$ -	$ -	$ -	$ -	$ 133,722	$ 137,733	$ 141,865	$ 146,121	$ 150,505	$ 155,020	$ 159,671	$ 164,461	$ 169,395	$ 174,476
Variable	$ -	$ -	$ -	$ -	$ -	$ 112,445	$ 115,819	$ 119,293	$ 122,872	$ 126,558	$ 130,355	$ 134,265	$ 138,293	$ 142,442	$ 146,716
Fuel	$ -	$ -	$ -	$ -	$ -	$ 240,475	$ 247,689	$ 255,120	$ 262,773	$ 270,657	$ 278,776	$ 287,140	$ 295,754	$ 304,626	$ 313,765
Operating income	$ -	$ -	$ -	$ -	$ -	$ 772,306	$ 795,476	$ 819,340	$ 843,920	$ 869,238	$ 895,315	$ 922,174	$ 949,839	$ 978,335	$ 1,007,685
Interest Expense	$ -	$ -	$ -	$ -	$ -	$ 161,335	$ 159,910	$ 158,372	$ 156,711	$ 154,917	$ 152,980	$ 150,887	$ 148,627	$ 146,186	$ 143,550
Depreciation & Amortisation	$ -	$ -	$ -	$ -	$ -	$ 150,055	$ 288,866	$ 267,178	$ 247,171	$ 228,604	$ 211,478	$ 195,592	$ 180,947	$ 178,546	$ 178,506
Taxable Income	$ -	$ -	$ -	$ -	$ -	$ 460,917	$ 346,699	$ 393,789	$ 440,038	$ 485,717	$ 530,858	$ 575,695	$ 620,266	$ 653,603	$ 685,629
Income Taxes	$ -	$ -	$ -	$ -	$ -	$ 175,148	$ 131,746	$ 149,640	$ 167,214	$ 184,572	$ 201,726	$ 218,764	$ 235,701	$ 248,369	$ 260,539
Net Income	$ -	$ -	$ -	$ -	$ -	$ 285,768	$ 214,953	$ 244,149	$ 272,824	$ 301,144	$ 329,132	$ 356,931	$ 384,565	$ 405,234	$ 425,090
Cash form Operation	$ -	$ -	$ -	$ -	$ -	$ 772,306	$ 795,476	$ 819,340	$ 843,920	$ 869,238	$ 895,315	$ 922,174	$ 949,839	$ 978,335	$ 1,007,685
Income Taxes	$ -	$ -	$ -	$ -	$ -	$ 175,148	$ 131,746	$ 149,640	$ 167,214	$ 184,572	$ 201,726	$ 218,764	$ 235,701	$ 248,369	$ 260,539
Total Interest Expense	$ -	$ -	$ -	$ -	$ -	$ 161,335	$ 159,910	$ 158,372	$ 156,711	$ 154,917	$ 152,980	$ 150,887	$ 148,627	$ 146,186	$ 143,550
Total Principal Repayment	$ -	$ -	$ -	$ -	$ -	$ 17,802	$ 19,226	$ 20,764	$ 22,426	$ 24,220	$ 26,157	$ 28,250	$ 30,510	$ 32,950	$ 35,587
Operating Cash Flow	$ -	$ -	$ -	$ -	$ -	$ 418,021	$ 484,593	$ 490,563	$ 497,569	$ 505,529	$ 514,452	$ 524,273	$ 535,002	$ 550,829	$ 568,009
Capital Cost	$ 328,217	$ 1,020,098	$ 880,685	$ 729,912	$ 567,141	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ (328,217)	$ (1,020,098)	$ (880,685)	$ (729,912)	$ (567,141)	$ 418,021	$ 484,593	$ 490,563	$ 497,569	$ 505,529	$ 514,452	$ 524,273	$ 535,002	$ 550,829	$ 568,009
Loan Draws	$ 164,108.50	$ 510,049.22	$ 440,342.49	$ 364,955.86	$ 283,570.70	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ (164,109)	$ (510,049)	$ (440,342)	$ (364,956)	$ (283,571)	$ 418,021	$ 484,593	$ 490,563	$ 497,569	$ 505,529	$ 514,452	$ 524,273	$ 535,002	$ 550,829	$ 568,009
Equity Draws	$ 164,109	$ 510,049	$ 440,342	$ 364,956	$ 283,571	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ -	$ -	$ -	$ -	$ -	$ 418,021	$ 484,593	$ 490,563	$ 497,569	$ 505,529	$ 514,452	$ 524,273	$ 535,002	$ 550,829	$ 568,009

Internal rate of Return		**20.00%**
Net Present Value at discount rate		
8%	$	3,332,799
10%	$	2,206,053
12%	$	1,432,667
14%	$	888,975
20%	$	0

Table A5-16. Cash flow analysis for the Sour PSA coal-to-methanol case, in $ x1000 (cont.).

end of	year 16	year 17	year 18	year 19	year 20	year 21	year 22	year 23	year 24	year 25	year 26	year 27	year 28	year 29	year 30	year 31	year 32	year 33	year 34	year 35
Operating revenues	$ 1,691,921	$ 1,742,679	$ 1,794,959	$ 1,848,808	$ 1,904,272	$ 1,961,400	$ 2,020,242	$ 2,080,850	$ 2,143,275	$ 2,207,573	$ 2,273,801	$ 2,342,015	$ 2,412,275	$ 2,484,643	$ 2,559,183	$ 2,635,958	$ 2,715,037	$ 2,796,488	$ 2,880,382	$ 2,966,794
Operating expenses																				
Fixed	$ 179,711	$ 185,102	$ 190,655	$ 196,375	$ 202,266	$ 208,334	$ 214,584	$ 221,022	$ 227,652	$ 234,482	$ 241,516	$ 248,762	$ 256,225	$ 263,911	$ 271,829	$ 279,983	$ 288,383	$ 297,034	$ 305,945	$ 315,124
Variable	$ 151,117	$ 155,651	$ 160,320	$ 165,130	$ 170,084	$ 175,186	$ 180,442	$ 185,855	$ 191,430	$ 197,173	$ 203,089	$ 209,181	$ 215,457	$ 221,920	$ 228,578	$ 235,435	$ 242,498	$ 249,773	$ 257,267	$ 264,985
Fuel	$ 323,178	$ 332,874	$ 342,860	$ 353,146	$ 363,740	$ 374,652	$ 385,892	$ 397,468	$ 409,392	$ 421,674	$ 434,324	$ 447,354	$ 460,775	$ 474,598	$ 488,836	$ 503,501	$ 518,606	$ 534,164	$ 550,189	$ 566,695
	$ 654,006	$ 673,626	$ 693,835	$ 714,650	$ 736,089	$ 758,172	$ 780,917	$ 804,345	$ 828,475	$ 853,329	$ 878,929	$ 905,297	$ 932,456	$ 960,430	$ 989,243	$ 1,018,920	$ 1,049,488	$ 1,080,972	$ 1,113,401	$ 1,146,803
Operating income	$ 1,037,915	$ 1,069,053	$ 1,101,124	$ 1,134,158	$ 1,168,183	$ 1,203,228	$ 1,239,325	$ 1,276,505	$ 1,314,800	$ 1,354,244	$ 1,394,871	$ 1,436,717	$ 1,479,819	$ 1,524,213	$ 1,569,940	$ 1,617,038	$ 1,665,549	$ 1,715,516	$ 1,766,981	$ 1,819,991
Interest Expense	$ 140,703	$ 137,629	$ 134,308	$ 130,722	$ 126,848	$ 122,665	$ 118,148	$ 113,268	$ 107,999	$ 102,308	$ 96,162	$ 89,524	$ 82,355	$ 74,612	$ 66,250	$ 57,219	$ 47,466	$ 36,932	$ 25,556	$ 13,269
Depreciation & Amortisation	$ 178,546	$ 178,506	$ 178,546	$ 178,506	$ 178,546	$ 178,506	$ 178,546	$ 178,506	$ 178,546	$ 178,506	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Taxable Income	$ 718,666	$ 752,919	$ 788,271	$ 824,931	$ 862,789	$ 902,057	$ 942,632	$ 984,731	$ 1,028,255	$ 1,073,430	$ 1,298,710	$ 1,347,194	$ 1,397,464	$ 1,449,601	$ 1,503,690	$ 1,559,819	$ 1,618,083	$ 1,678,583	$ 1,741,425	$ 1,806,721
Income Taxes	$ 273,093	$ 286,109	$ 299,543	$ 313,474	$ 327,860	$ 342,782	$ 358,200	$ 374,198	$ 390,737	$ 407,904	$ 493,510	$ 511,934	$ 531,036	$ 550,849	$ 571,402	$ 592,731	$ 614,872	$ 637,862	$ 661,742	$ 686,554
Net Income	$ 445,573	$ 466,809	$ 488,728	$ 511,457	$ 534,929	$ 559,275	$ 584,432	$ 610,533	$ 637,518	$ 665,527	$ 805,200	$ 835,260	$ 866,428	$ 898,753	$ 932,288	$ 967,088	$ 1,003,212	$ 1,040,722	$ 1,079,684	$ 1,120,167
Cash form Operation	$ 1,037,915	$ 1,069,053	$ 1,101,124	$ 1,134,158	$ 1,168,183	$ 1,203,228	$ 1,239,325	$ 1,276,505	$ 1,314,800	$ 1,354,244	$ 1,394,871	$ 1,436,717	$ 1,479,819	$ 1,524,213	$ 1,569,940	$ 1,617,038	$ 1,665,549	$ 1,715,516	$ 1,766,981	$ 1,819,991
Income Taxes	$ 273,093	$ 286,109	$ 299,543	$ 313,474	$ 327,860	$ 342,782	$ 358,200	$ 374,198	$ 390,737	$ 407,904	$ 493,510	$ 511,934	$ 531,036	$ 550,849	$ 571,402	$ 592,731	$ 614,872	$ 637,862	$ 661,742	$ 686,554
Total Interest Expense	$ 140,703	$ 137,629	$ 134,308	$ 130,722	$ 126,848	$ 122,665	$ 118,148	$ 113,268	$ 107,999	$ 102,308	$ 96,162	$ 89,524	$ 82,355	$ 74,612	$ 66,250	$ 57,219	$ 47,466	$ 36,932	$ 25,556	$ 13,269
Total Principal Repayment	$ 38,433	$ 41,508	$ 44,829	$ 48,415	$ 52,288	$ 56,471	$ 60,989	$ 65,868	$ 71,138	$ 76,829	$ 82,975	$ 89,613	$ 96,782	$ 104,525	$ 112,886	$ 121,917	$ 131,671	$ 142,204	$ 153,581	$ 165,867
Operating Cash Flow	$ 585,685	$ 603,807	$ 622,445	$ 641,548	$ 661,186	$ 681,310	$ 701,988	$ 723,170	$ 744,926	$ 767,204	$ 722,225	$ 745,647	$ 769,646	$ 794,228	$ 819,401	$ 845,170	$ 871,541	$ 898,517	$ 926,103	$ 954,300
Capital Cost	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Investment	$ 585,685	$ 603,807	$ 622,445	$ 641,548	$ 661,186	$ 681,310	$ 701,988	$ 723,170	$ 744,926	$ 767,204	$ 722,225	$ 745,647	$ 769,646	$ 794,228	$ 819,401	$ 845,170	$ 871,541	$ 898,517	$ 926,103	$ 954,300
Loan Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow after Debt Financing	$ 585,685	$ 603,807	$ 622,445	$ 641,548	$ 661,186	$ 681,310	$ 701,988	$ 723,170	$ 744,926	$ 767,204	$ 722,225	$ 745,647	$ 769,646	$ 794,228	$ 819,401	$ 845,170	$ 871,541	$ 898,517	$ 926,103	$ 954,300
Equity Draws	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -	$ -
Net Cash Flow for Equity Distribution	$ 585,685	$ 603,807	$ 622,445	$ 641,548	$ 661,186	$ 681,310	$ 701,988	$ 723,170	$ 744,926	$ 767,204	$ 722,225	$ 745,647	$ 769,646	$ 794,228	$ 819,401	$ 845,170	$ 871,541	$ 898,517	$ 926,103	$ 954,300