# Final Report

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***Project Title:*** *Advanced Acid Gas Separation Technology for the Utilization of Low Rank Coals*

***Project Period:*** *October 1, 2011 to December 31, 2012*

**Date of Report:** December 2012

**Recipient:** Air Products & Chemicals, Inc.
**Duns Number:** 003001070
**Award Number:** DE–FE0007759

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**Acknowledgment:** “This material is based upon work supported by the Department of Energy under Award Number DE-FE0007759.”

**Abstract**: 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 with higher rank coals to provide a significant reduction in the cost of $CO_2$ capture for power generation, which should translate to a reduction in cost of electricity (COE), compared to baseline $CO_2$ capture plant design. The objective of this project is to test the performance and capability of the adsorbents in handling tar and other impurities using a gaseous mixture generated from the gasification of lower rank, lignite coal. The results of this testing are used to generate a high-level pilot process design, and to prepare a techno-economic assessment evaluating the applicability of the technology to plants utilizing these coals.

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# Introduction: Advanced Acid Gas Separation Technology for the Utilization of Low Rank Coals

## 1- 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 with higher rank coals to provide a significant reduction in the cost of $CO_2$ capture for power generation, which should translate to a reduction in cost of electricity (COE), compared to baseline $CO_2$ capture plant design. The objective of this project is to test the performance and capability of the adsorbents in handling tar and other impurities using a gaseous mixture generated from the gasification of lower rank, lignite coal. The results of this testing are used to generate a high-level pilot process design, and to prepare a techno-economic assessment evaluating the applicability of the technology to plants utilizing these coals.

## 2- Background:

Gasification is a promising alternative to traditional coal-fired combustion processes for power production that can also be adapted to supply hydrogen or liquid fuel precursors, thereby expanding coal markets. The key challenge for gasification is to reduce its cost. This is especially true for clean coal projects with significant levels of $CO_2$ capture. The conventional technology option for capturing $CO_2$ from gasification processes involves traditional AGR technologies, based on a physical solvent adsorption process that utilizes Selexol™ or Rectisol® to selectively separate $H_2S$ and $CO_2$ from $H_2$ and CO. This process is sizeable portion of the plant capital and requires significant utility consumption during operation, which only escalates with increasing levels of $CO_2$ capture. Air Products has developed an alternative AGR technology, Sour PSA, which removes $H_2S$ and $CO_2$ from gasification streams utilizing a solid adsorbent-based process. When configured as a capture option, this process can achieve a higher level of $CO_2$ capture than conventional technology at significantly lower capital and operating costs, and is expected to reduce the cost of $CO_2$ capture by over 20%.

Air Products developed this novel technology by leveraging years of experience in the design and operation of $H_2$ PSA systems in Air Products' numerous steam methane reformers (SMRs). For application to gasification, Air Products' commercial PSAs were designed to operate on clean, sweet synthesis gas (syngas) and thus required an upstream AGR unit. Air Products has extensive experience with $H_2$ PSA technology. This served as a foundation for development of the Sour PSA concept for handling sour syngas feed streams, thereby eliminating the need for the upstream AGR unit.

The system consists of three components: i) a PSA unit for purification of the sour syngas stream; ii) a sulfur treatment unit, and iii) a $CO_2$ polishing and compression unit. Products including high-purity $H_2$, low-purity $H_2$, and $CO/H_2$ synthesis gas mixtures can be purified at pressure by the Sour PSA system. By combining the impurities stream of Sour PSA with a tailored tail gas treatment technology, a purified, capture-grade $CO_2$ stream can be produced for use in enhanced oil recovery or terrestrial sequestration. The impurities in the $CO_2$ stream are either further processed into a desired form, or recycled back within the process. By-product $H_2S$ can then be recovered either as sulfuric acid, as a neutralized, acid-derived product like gypsum, or in elemental form (e.g., Claus process). Product molecules can be recycled back into the process to improve overall plant efficiencies, while inert-containing streams with compounds like Ar and $N_2$ are treated and vented.

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 and 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.

Traditional AGR technologies are known to be effective for the removal of additional minor contaminants in the syngas. Compounds such as heavier hydrocarbons and acidic gases are readily soluble in the solvents used in commercial AGR adsorption technologies like Rectisol® and Selexol™. Air Products' Sour PSA process must be designed to be at least as robust as current technologies if it is to be a commercially viable option. This requires data based on exposure of the adsorbent to impurities of real gasifier-derived syngas. The final product specifications are equally important when determining which impurities in the syngas need to be removed before further processing occurs. If producing power, these purification steps will be designed to meet emission regulations at the stack of the plant or, in a clean power case, $CO_2$ capture requirements. In a high-purity $H_2$ application, such as refinery $H_2$, reducing CO, $N_2$, and Ar levels to parts per million (ppm) may be required. Additional challenges arise when purifying mixtures of CO and $H_2$ for further processing into chemicals like Fischer Tropsch products and methanol or gases like substitute natural gas (SNG). These processes require a carefully controlled, optimal ratio of CO and $H_2$. This is a challenge for today's current commercial technologies, as well as Air Products' Sour PSA technology.

## 3- Approach:

The project has been conducted in a single phase consisting of five tasks, the first task being the overarching effective management of the project.

Air Products' existing Sour PSA test unit has been modified to improve operability and reliability (Task 2). The test unit has then been attached to EERC's fluidized bed gasifier and operated utilizing shifted syngas derived from oxy-blown gasification of Montana Rosebud PRB coal in both PSA and TSA modes to determine the suitability of the unit to operate with such coal (Task 3).

The experimental results have been used to provide information necessary for a high level process design for a non site specific Sour PSA pilot unit (Task 4).

Simulations based on the experimental results are used to estimate full scale equipment size and performance in order to build a techno-economic assessment (Task 5).

Ultimately, methodology reported in NETL's "Cost and Performance Baseline for Fossil Energy Plants: low rank coal to electricity: IGCC cases", has been used to provide direct comparison of the resulting Cost Of Electricity from IGCC plant using traditional AGR technologies

# Task 2. Modify PSA/TSA unit for tar gas operations and install at EERC:

The equipment used in this work consisted of a sour PSA/TSA unit, a pilot-scale fluidized bed gasifier, and assorted analytical instruments.

## 1- Sour PSA/TSA Unit

The sour PSA/TSA unit was previously built at Air Products and Chemicals, Inc (APCI), and was used at the Energy and Environmental Research Center (EERC) facility to test the performance of various adsorbent materials under PSA or TSA conditions while feeding with sour syngas from the gasifier. It consisted of two insulated and electrically heated columns, a set of air-actuated switching valves at the top and bottom of the column, feed and purge gas mass flow controllers, various process tanks (product, depressurization, and purge), effluent gas flow meters, and analytical detectors (micro GC and CO IR unit). A schematic and photograph of the unit is illustrated in Figures T2-1 and T2-2.

The system could be operated under a PSA or TSA cycle. In PSA mode each bed sequentially progresses through the following process steps;

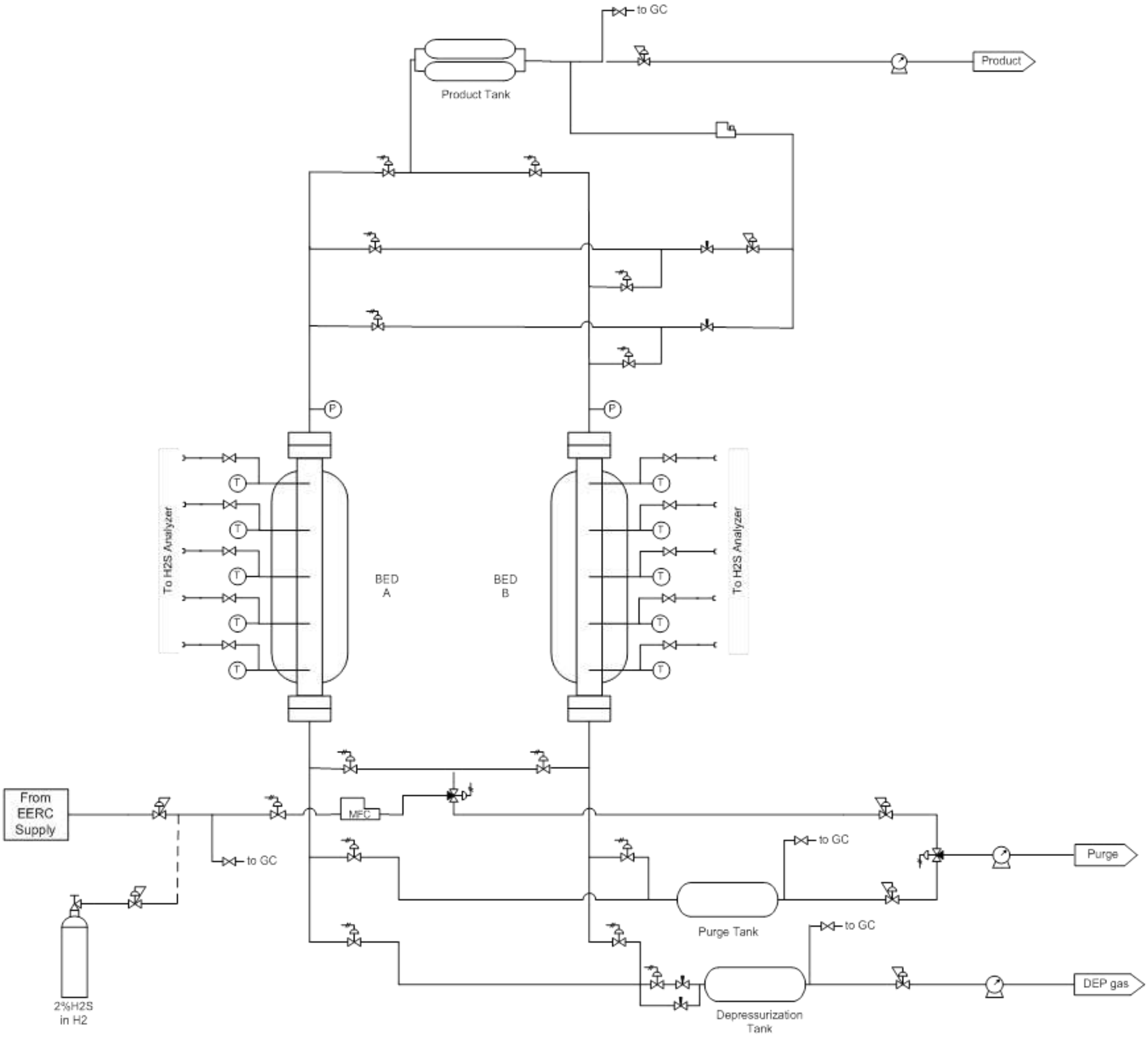
1. High pressure sour syngas feed to the bed (normally 400 psig)
2. Countercurrent depressurization of the bed (to ~1-10 psig)
3. Countercurrent low pressure purge of the bed (1-5 psig) with product gas
4. Countercurrent re-pressurization of the bed with product gas.

In the case of TSA operation, the low pressure purge step was subdivided into heating, holding, and cooling steps.

The system was operated through a Programmable Logic Controller (PLC) system which automatically executed the proper valve and heater sequencing for the chosen cycle. The operator set the time lengths of the various steps, the feed flow rate, the purge gas flow rate (via a manual purge valve), the repress rate (via a manual repress valve), and the depressurization gas flow rate (via a set of parallel manual depress valves). Gas pressure during each step was controlled by back pressure regulators. Gas flow rates and compositions were determined for the feed, product, purge effluent and depress effluent streams. This information was then used to evaluate overall and component mass balances and estimate the $H_2$ recovery/$H_2S$ rejection.

An Inficon micro Gas Chromatograph (micro GC) unit was used for gas analysis, and it was capable of monitoring all of the typical syngas species as well as $H_2S$ and COS. A second micro-GC (referred to as the 'tar GC') with a wax column was installed for analysis of tar species, especially benzene and substituted aromatics.

The Sour PSA/TSA unit was also used to measure breakthrough of $H_2S$, $CO_2$ or CO in an inert carrier gas ($H_2$ or He). In this case, the feed gas was supplied from a gas cylinder and a fixed flow rate was delivered to each packed column with a mass flow controller. The adsorbent was previously regenerated by purging with $N_2$ and pressurized to 400 psig in $H_2$. 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 at a mid-bed port with the micro-GC. Flow continued until the concentration of the contaminant gas in the effluent stream was roughly 60% of the feed gas composition.



FigureT2-1. Schematic of Sour PSA/TSA Unit.

Figure T2-2. Photo of the Sour PSA/TSA Unit.

Significant modifications were made to the original PSA/TSA unit for this study. This included installation of the ‘tar GC’, a Coriolis flow meter for the feed gas, seven pressure transmitters, multiple air actuated and manual valves, and four electronic needle valve positioners. In addition, the CO, $H_2S$ and flammable gas monitors and ventilation flow switch assembly were replaced with explosion-proof versions. An industrial analyzer cabinet was obtained to house the two GC units and $H_2S$ Continuous Emission Monitor (CEM). Industrial Z-purge and X-purge systems were installed on the PSA unit electronics cabinet and the analyzer cabinet, respectively. Relief valves on the system were all removed, cleaned, and recertified. A series of new calibration gases were obtained for the GC unit.

The safety philosophy for the unit was modified since the system was located in an electrically classified area. Safety analysis suggested that pressurizing and purging the entire PSA enclosure with $N_2$ would be the optimum approach. This required sealing the existing system enclosure and developing an approach for supplying adequate amounts of house N2 to the system. The sealing task was quite significant, as the unit was initially designed for ventilation rather than pressurization. It required installation of a solid base for the unit, patching of various piping/access holes, and caulking of seams in the enclosure. House $N_2$ was piped to a distributor assembly installed at the bottom of the enclosure, and adequate valving and flow meters were included to assist with setting the appropriate purge flow rate. An industrial pressure switch was installed to shut down the system if positive pressure in the enclosure was lost.

Safety review meetings were completed at APCI in early February. The unit was shipped to EERC's National Center for Hydrogen Technologies facility in Grand Forks, ND on 15 February. Installation of the unit along with auxiliary equipment (gas cylinders, analyzers, sample/vent lines) was conducted by APCI and EERC during the following two weeks.

In Campaign 2 (described later), a relatively minor plumbing modification was made to minimize gas burping to the column during the start of the PSA purge step. An automated shutoff valve was added immediately after the purge gas needle valve which closed during all steps other than the purge, preventing buildup of gas in the supply line and subsequent burping.

A simple six-bed exposure unit was built in this program to investigate the adsorption of tar components on up to six different adsorbents. A photo of the parallel beds associated with that unit is illustrated in Figure T2-3. The beds were made from 3/8" steel tubing which were loaded with a few grams of crushed and sieved adsorbent material. A gas manifold assembly was interfaced to the beds to allow a 1-2 lpm flow of sour syngas to flow through the beds at ambient temperature and ~400 psig. This flow continued for a number of days, with the goal of fully loading the samples with the trace tar species. The column assembly was then removed, capped, and sent to APCI for analysis.

Figure T2-3. Photo of the Six-bed Sour Syngas Exposure Assembly

## 2- EERC Gasifiers

The Sour PSA/TSA test unit was interfaced with a pilot scale fluidized bed gasifier (FBG) at EERC. It was housed in high bay space at the EERC's National Center for Hydrogen Technology and was commissioned around 2009. Coal crushed to -10 mesh is the feedstock for this unit, and it is fed to the unit without any pre-drying. It is combined with a bed of silica sand fluidized with a combination of steam/$O_2$ and recycle syngas along with a small amount of nitrogen from the coal feeder purges. Sand is used to maintain the reactor temperature and transfer energy to the incoming coal. The coal is gasified in the reactor, and a sour syngas is recovered after rejecting solid ash from the gaseous reactor effluent.

This system was designed according to American Society of Mechanical Engineers (ASME) B31.3 Process Piping Code specifications. The reactor was designed with the capability to operate at a MOP of 1000 psig at an operational temperature of 843°C, 650 psig at an operational temperature of 917°C, and 300 psig at an operational temperature of 982°C. The system was electrically heated via external heating circuits. Haynes 556® alloy was selected as the material of construction for the reactor, all the reactor nozzles, and the cyclone. A design drawing of the reactor is shown in Figure T2-4, and a photograph of the gasifier is shown in Figure T2-5. A design drawing of the fuel feed system is shown in Figure T2-6, with a photograph of the feeder vessel in Figure T2-7.

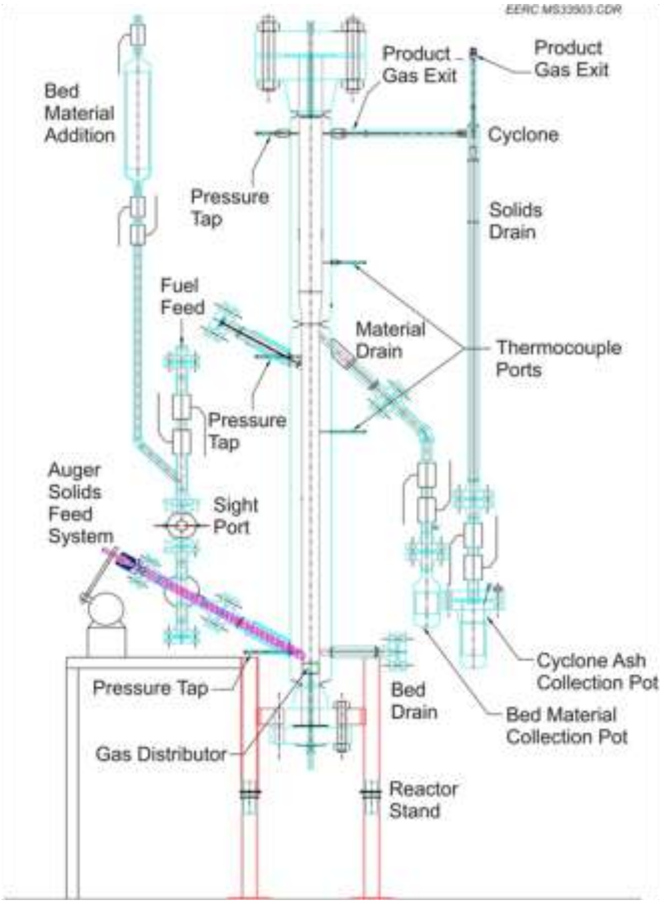


Figure T2-4. Design drawing of the pressurized, fluidized gasification reactor.


EERC MS34917.CDR

Figure T2-5. Photograph of the high-pressure FBG.

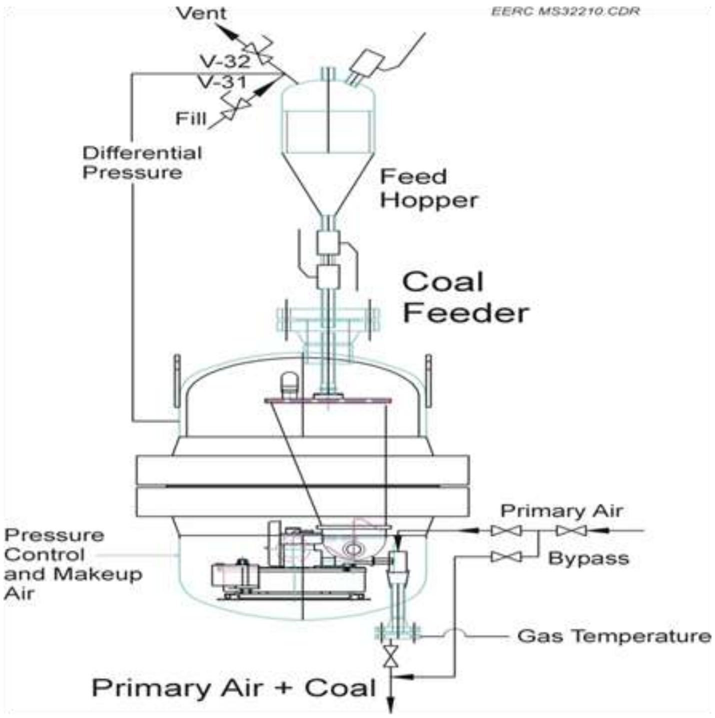

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Figure T2-6. Cross-sectional view of the fuel feed system.

Figure T2-7. Photograph of the fluid-bed coal feed system.

The feed system uses a K-tron® loss-in-weight feeder inside a pressure vessel capable of 1000 psig operation. This system permits real-time measurement of the fuel feed rate to the gasification system. The feed system electronic controls are interfaced to a data acquisition system that allows for local or remote computer control of the fuel feed rate. The upper pressure vessel is the fuel charge hopper. The fuel charge hopper is manually charged with fuel through the top valve while at atmospheric pressure. It is then sealed and pressurized. Finally, the fuel feed material is transferred by gravity feed to the weigh hopper inside through the lower dual-valve system. The weigh hopper is on an integral platform scale that provides an electronic signal of the overall weight of the fuel feed material. Hopper weights along with feed rates are recorded by the data acquisition system and can be displayed and trended as required.

After ash removal, the syngas is passed to a set of four fixed bed reactors rated for operation at 540°C and 1000 psig. These fixed bed reactors are approximately 5 inches in diameter and 26 inches long and are electrically heated. They can be operated in series or parallel configuration. In this work, they were packed with sour shift catalysts and used to shift CO and water to more $H_2$ and $CO_2$. Two sets of three (six total) water-cooled quench pots were located after the reactors to cool the shifted syngas and condense moisture and organics from the gas stream. The quench pots were designed for operation up to 1000 psig. Either water or a cooled glycol and water mixture was circulated through the outer jacket of each quench pot to cool the product gas.

### 3- Analytical testing at APCI

To gauge adsorbent stability, $CO_2$ capacities for the fresh and spent samples were determined by thermal gravimetric adsorption (TGA) analysis. The TGA analyzer (TA Instruments Q5000) consists of a very sensitive balance capable of measuring the mass change of an adsorbent sample under various gas atmospheres and temperatures. The standard TGA procedure used to determine $CO_2$ capacity involved heating the sample to 200$^o$C under $N_2$ purge, cooling the sample to 40$^o$C, and exposing it to 100% dry $CO_2$ at 1 atmosphere. The final steady weight change yielded a measure of the $CO_2$ adsorption capacity. A Dycor System 2000 mass spectrometer attached to the TGA furnace exit was used to identify species evolved upon heating to 200$^o$C.

Conventional low temperature $N_2$ adsorption techniques were used to quantify the adsorbent surface area and provide details on the pore volume of the samples (conducted after an initial regeneration under vacuum at 200° C).

Elemental compositions of the adsorbent samples were determined by x-ray fluorescence (XRF).

A combination of gas chromatography mass spectroscopy (GCMS) and headspace analysis was conducted with some adsorbent samples to determine what species were adsorbed. Samples for GCMS analysis were heated in a $N_2$ purged glass tube to 200$^o$C and evolved gases were collected on a Tenax tube. The tube was thermally desorbed onto a 50-meter Agilent Ultra 2 capillary column using a Gerstel Thermal Desorption System and Cooled Inlet System. Analysis was carried out using an Agilent mass selective detector.

# Task 3. Run Sour PSA/TSA unit with shifted sour syngas derived from EERC fluidized bed gasifier with a Montana Rosebud PRB coal in oxy-blown mode:

The objective of the experimental phase of this project was to test the performance and capability of the sour PSA process in handling tar and other impurities from the gasification of PRB coal. Namely, cyclic PSA tests were conducted to evaluate process performance. Multiple adsorbents were exposed to the tar-loaded gas and tested to screen potential applicability in the PSA or a TSA process. The last fixed bed test focused on the performance of some of these materials in a TSA cycle. The experimental data were then used to validate an internal adsorption process simulator which was subsequently used for industrial process design.

Experimental testing of the EERC FBG / Sour PSA/TSA system was conducted between February – June 2012. The first one week test (Campaign 1) was conducted 4 - 9 March 2012. The goal was to identify any operational issues with the PSA or gasifier, and get some run time on the PSA adsorbent. A second two week run (Campaign 2) ran from 9 - 19 April 2012 and focused on long term operation and the ability to achieve a cyclic steady state (CCS) with the PSA. A final one week test (Campaign 3) held on 20-25 May 2012 investigated the effectiveness of TSA cycles with the same sour syngas.

In all of these tests, Montana Rosebud PRB coal was gasified in EERC's pilot-scale fluidized bed gasifier under oxygen-blown conditions to create a sour syngas. This gas was passed through a sour water gas shift bed to convert most of the CO to $CO_2$. The bulk of the condensables were then removed in a water-based quench pot. The cooled, shifted sour syngas was fed to the sour PSA/TSA unit.

An analysis of the coal feedstock is provided in Table T3-1. The as-received coal was fed to the fluidized bed gasifier; no additional drying was required.

The various adsorbents utilized in this work are listed below in Table T3-2. These materials are all commercially available and were obtained directly from the vendors without any pretreatment or modification.

In addition to the fixed bed testing with the PSA/TSA unit, the six-bed exposure system was used to expose various adsorbents to the raw, sour syngas. Adsorbent samples were crushed and sieved to 20/30 mesh size and exposed to flowing syngas at ~400 psig and ambient temperature for multiple days. The samples were then extracted and analyzed in APCI labs.

## 1- Results from Campaign 1

Two PSA vessels were each packed with 87 inches of ADS_A and 6 inches of ADS_C, with the ADS_C located at the feed end of the column. They were installed into the sour PSA/TSA unit and regenerated under flowing $N_2$ at 5 psig and 150$^{o}$C for 4 hrs.

## Table T3-1. Analysis of Montana Rosebud PRB Coal.

| | As Det. (%) | As Recd. (%) | Dry (%) | Dry/Ash Free (%) |
|---|---|---|---|---|
| | **Proximate Analysis** | | | |
| **Moisture** | 6.75 | 25.77 | N/A | N/A |
| **Volatile Matter** | 34.83 | 27.72 | 37.35 | 40.05 |
| **Fixed Carbon (I nd)** | 52.13 | 41.49 | 55.90 | 59.95 |
| **Ash** | 6.29 | 5.01 | 6.75 | N/A |
| | | | | |
| | **Ultimate Analysis** | | | |
| **Hydrogen** | 4.92 | 6.20 | 4.47 | 4.79 |
| **Carbon** | 63.47 | 50.52 | 68.06 | 72.99 |
| **Nitrogen** | 0.95 | 0.76 | 1.02 | 1.09 |
| **Sulfur** | 0.98 | 0.78 | 1.05 | 1.13 |
| **Oxygen (Ind)** | 23.38 | 36.73 | 18.65 | 20.00 |
| **Ash** | 6.29 | 5.01 | 6.75 | N/A |
| | | | | |
| **Heating Value (Btu/lb)** | 10581 | 8422 | 11347 | 12168 |

| Oxides (wt.%) | (a) | (b) | (c) | Elemental (wt.%) | (d) | (e) |
|---|---|---|---|---|---|---|
| $SiO_2$ | 39.92 | 39.92 | 46.00 | **Si** | 35.29 | 39.22 |
| $Al_2O_3$ | 18.61 | 18.61 | 21.44 | **Al** | 18.63 | 20.70 |
| $Fe_2O_3$ | 6.25 | 6.25 | 7.20 | **Fe** | 8.27 | 9.19 |
| $TiO_2$ | 1.32 | 1.32 | 1.52 | **Ti** | 1.50 | 1.66 |
| $P_2O_5$ | 0.31 | 0.31 | 0.36 | **P** | 0.26 | 0.28 |
| **CaO** | 12.01 | 12.01 | 13.84 | **Ca** | 16.23 | 18.04 |
| **MgO** | 7.53 | 7.53 | 8.68 | **Mg** | 8.59 | 9.55 |
| $Na_2O$ | 0.51 | 0.51 | 0.58 | **Na** | 0.71 | 0.79 |
| $K_2O$ | 0.33 | 0.33 | 0.38 | **K** | 0.52 | 0.58 |
| $SO_3$ | 13.21 | 13.21 | ----- | **S** | 10.01 | ----- |

(a) Oxide concentrations (wt.%) on an ash basis.
(b) Oxide concentrations normalized to a closure of 100%.
(c) Oxide concentrations renormalized to a SO3-free basis.
(d) Elemental concentrations (wt.%) on an ash basis.
(e) Elemental concentrations renormalized to a S-free basis.

**Table T3-2. Adsorbents used in this work.**

| Acronym |
|---|
| ADS_A |
| ADS_B |
| ADS_C |
| ADS_D |
| ADS_E |
| ADS_F |
| ADS_G |
| ADS_H |

Sour syngas flow was started at 7 AM on 6 March and was terminated at 11:40 AM on 9 March. This gave a total on-stream time of 77 hours for the PSA unit. Both the gasifier and the PSA unit operated smoothly for this test. There were no unplanned shutdowns for either system. The PSA process cycle consisted of 1) 5 minutes of high pressure feed, 2) 2 minutes of countercurrent depressurization to ~10-20 psig, 3) 1 minute of low pressure countercurrent purge with a portion of the product gas at ~5-10 psig, and 4) two minutes of countercurrent re-pressurization to 400 psig with a second portion of the product gas.

The composition of the feed gas delivered to the PSA unit remained steady through the entire campaign. EERC continuously analyzed the feed gas to the PSA via multiple gas chromatographs. The largest deviation from the mean concentration for each species was 10% or less, with no overall trend in the deviations. The EERC composition data compared well with intermittent feed gas analyses obtained with the micro GC contained in the sour PSA system. Typical raw feed gas composition determined by our micro GC was 33.1% $H_2$, 7.7% $N_2$, 0.29% CO, 54.5% $CO_2$, 4.0% $CH_4$, 0.07% $C_2H_6$, and 0.295% $H_2S$. Trace tar components could not be determined by the available GC's (the tar GC was not available yet). A gas sample of the feed was withdrawn, though, and subsequent lab analysis (Intertek, attached in Appendix) indicated the presence of approximately 300 ppmv of $C_{6+}$ species, primarily benzene. In addition to the $H_2S$, lab analysis also indicated the presence of about 24 ppm of COS and 4 ppm of organic sulfur in the sour syngas.

The primary goal of Campaign 1 (and Campaign 2) was to characterize PSA performance at cyclic steady state (CSS), where the composition and flow of the PSA effluent streams becomes constant on a cycle-to-cycle basis. This point is typically reached by fixing two key PSA process parameters, the feed gas flow rate and purge rate, and monitoring the product gas composition. At cyclic steady state, the average product gas composition (measured after the product tank) will stabilize and remain constant with additional cycling.

A total of 453 PSA cycles were completed in this campaign. Of these, the majority of the first 179 cycles were conducted with lower purge flow. This quickly loaded the column with $H_2S$ and $CO_2$. The remaining 274 cycles were conducted at fixed feed and purge flow rates.

Product gas compositions obtained via the micro GC are plotted in Figure T3-1. Also shown at the top of the figure is a run timeline annotated with some of the more significant run notes. The results show the effect of the lower purge rate on the first day of operation – namely, loading of the column with $CO_2$ and $H_2S$ and a

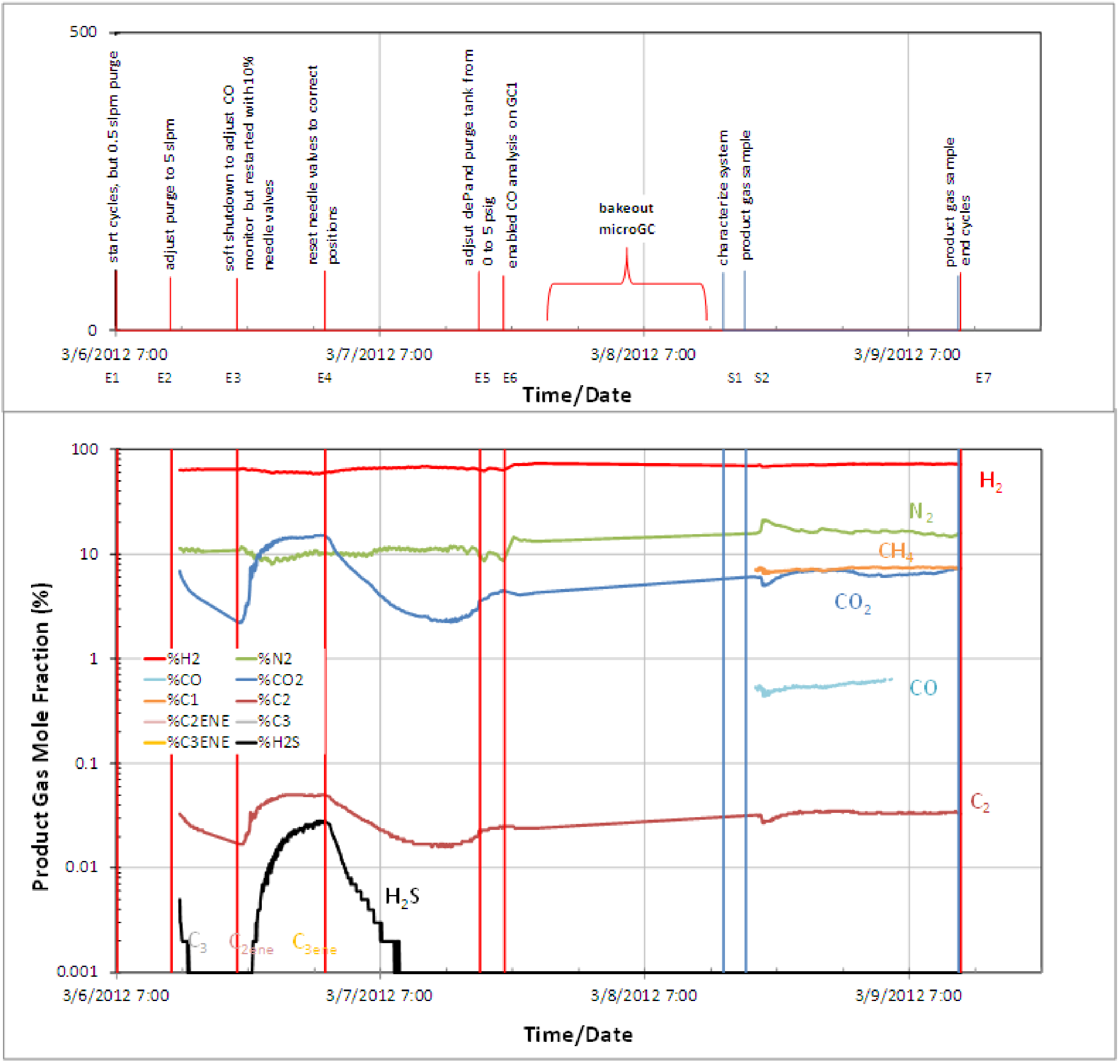


Figure T3-1. Product Gas Mole Fraction during Campaign 1.

strong increase in their concentrations in the product stream. Once the purge flow was set properly, though, the last 1/3 of the plot suggests that the $H_2$, $N_2$, $CO_2$, and $CH_4$ concentrations tend to stabilize while the $H_2S$ concentration drops to below the detection limit of the micro-GC (< 10 ppm). This low $H_2S$ level could be determined via Draeger tubes, and it ranged from about 4 to 8 ppm during this test. The scatter in the readings made it difficult to determine the trend in $H_2S$ level with time, but the last two data points were in the 8 ppm range suggesting that it was increasing.

The total gas volume and average concentration of the product gas, depressurization gas and purge gas were measured at around cycle 324. Combined with the feed gas flow rate and composition, this permitted evaluation of component mass balances and PSA performance estimations. These data are listed below in Table T3-3.

**Table T3-3. Performance data for the sour PSA at cycle 324, Campaign 1.**

| | **Mass Balance; In/Out** | | |
|---|---|---|---|
| **Total** | 1.02 | | |
| **$H_2$** | 1.03 | | |
| **$N_2$** | 1.09 | **$H_2$ recovery** | 54.5% |
| **CO** | 1.02 | **$H_2S$ Rejection** | 99.96% |
| **$CO_2$** | 1.03 | **$CO_2$ Rejection** | 96.9% |
| **$CH_4$** | 1.02 | **$H_2S$ in Product** | 5 ppm |
| **$C_2H_6$** | 1.03 | | |
| **$H_2S$** | 1.02 | | |

Component mass balances were generally within 3% ($N_2$ seems to be an outlier) which confirmed the integrity of the unit. The product gas consisted of 71% $H_2$, <7% $CO_2$, 5 ppm $H_2S$ (determined with Draeger tubes), along with $N_2$, CO, and $CH_4$ that pass through the adsorbent column. Thus, the PSA removes 99.96% of the incoming $H_2S$, from about 3000 ppm down to 5 ppm. It also rejects about 97% of the $CO_2$ in the feed stream. A sample of product gas was taken and shown to contain 6 ppm $H_2S$ and 0.7 ppm COS via lab analysis, which agrees well with the Draeger analysis. The $H_2$ recovery was calculated to be 55%. This is in line with expectations given the simple PSA cycle used. Addition of pressure equalization steps in an industrial unit will improve the recovery substantially.

More detailed investigation is needed to determine if the system is truly at CSS. An enlarged view of the product gas $CO_2$ and CO mole fraction at the end of the campaign is plotted in Figure T3-2. Both show a slow trend towards higher concentrations, on the order of 1 percentage point per 16 hours for $CO_2$ and 1 percentage point per 80 hrs for CO. Although these are slow increases, they indicate that both species are building up in the product gas and the PSA is not truly at cyclic steady state. A longer series of PSA cycles was necessary to evaluate this point further, and this was the focus of Campaign 2.

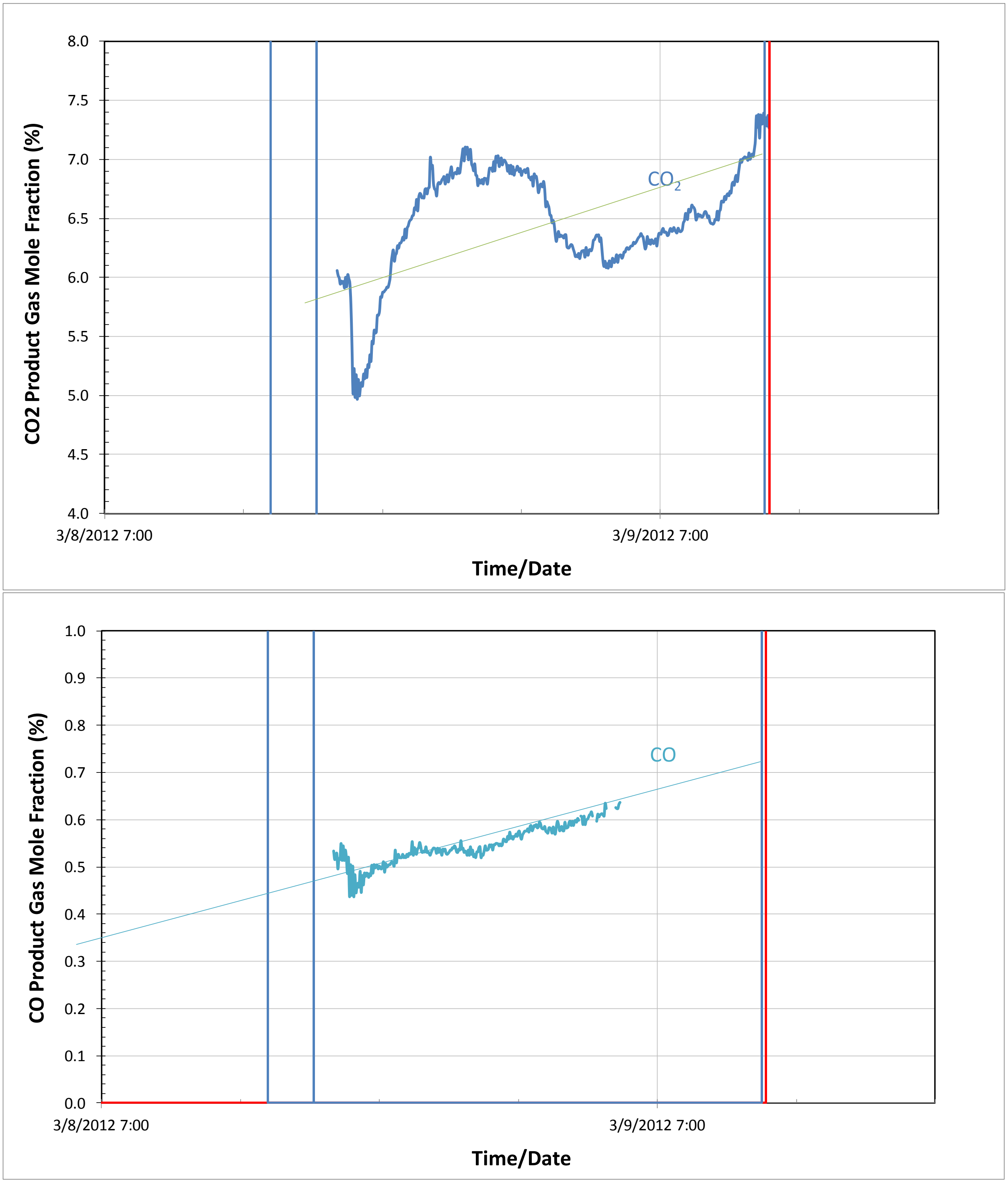


Figure T3-2. $CO_2$ and CO Mole Fractions in the Product Gas, Campaign 1.

## 2- Results from Campaign 2

This test was conducted over an extended two week period (9 to 19 April) to provide more time to achieve CSS. The PSA and gasifier were operated for roughly 200 hrs, yielding a total of 1193 PSA cycles. There was one major shutdown (12 April) during the test associated with an electrical outage at the facility which lasted for 18.5 hours.

The average raw feed gas composition determined by the micro GC was 32.2% $H_2$,6.2% $N_2$, 1.6% CO, 56.4% $CO_2$, 3.8% $CH_4$, 0.07% $C_2H_6$, and 0.31% $H_2S$. This is similar to the composition measured in Campaign 1 with the exception of CO, which increased from 0.3% to 1.6%. This is likely due to loss of water gas shift catalyst activity in the sour shift unit, or a change in operating conditions (e.g., temperature, steam/gas ratio) of that unit.

Constant feed gas, both in flow rate and composition, is a prerequisite to achieving CSS. Stability of the feed gas composition was determined from the EERC GC data. As in campaign 1, there is reasonably good agreement in trend and magnitude between feed gas compositions determined from the PSA micro GC and the various EERC GC's. Normalized residuals, defined as the difference between a given mole fraction and its average over the entire campaign all divided by the average concentration, were evaluated from the EERC data and are plotted in Figure T3-3. Nitrogen shows some significant spikes due to minor gasifier upsets and their associated syngas depletion/$N_2$ purging. Hydrogen and $CO_2$ residuals were within 10%, and $CH_4$ residuals were within 20%. CO varies quite a bit, presumably due to the previously mentioned sour shift activity changes or steam to syngas ratio variations enacted by EERC. The feed gas $H_2S$ content initially shows a declining trend, but after the gasifier shutdown/restart settles down at around 2600 ppm.

Information regarding the PSA cycling can be inferred from the plots in Figures T3-4 and T3-5. The pressure in Bed A and Bed B at the end of the feed step and the end of the purge step are plotted in Figure T3-4. The feed pressure ranges from 380 to 395 psig. The purge pressure was 1-2 psig higher for bed B than bed A, increased with purge gas flow rate, and overall ranged from 5.5-7.5 psig (bed A) and 7-9.5 psig (bed B). These are reasonable pressure points for the PSA. The feed and purge gas flow rates are plotted in Figure T3-5. These signals are taken as the analog output signal from the mass flow controller and meter. The feed flow rate was constant at 16 slpm during the entire run. The purge rate is one of the key PSA performance variables and was incrementally increased during the test. Measured gas flow rates in the effluent streams (product, purge, and depressurization outlets) are also plotted in Figure T3-5. The product DTM was continuously utilized; the purge and depressurization DTM's were typically bypassed to minimize corrosion of the meters. This yields intermittent flow readings for the purge and depressurization streams. The flow rates are reasonably consistent.

Complete mass balance measurements were conducted during two points of operation as illustrated in Figure T3-6. The overall mass balance is excellent, and component mass balances (excluding CO which could not be calculated due to GC issues) were typically within 10%. In both cases, the $H_2$ recovery is about 59%, $H_2S$ rejection is 99.95%, and $CO_2$ rejection is 96%.

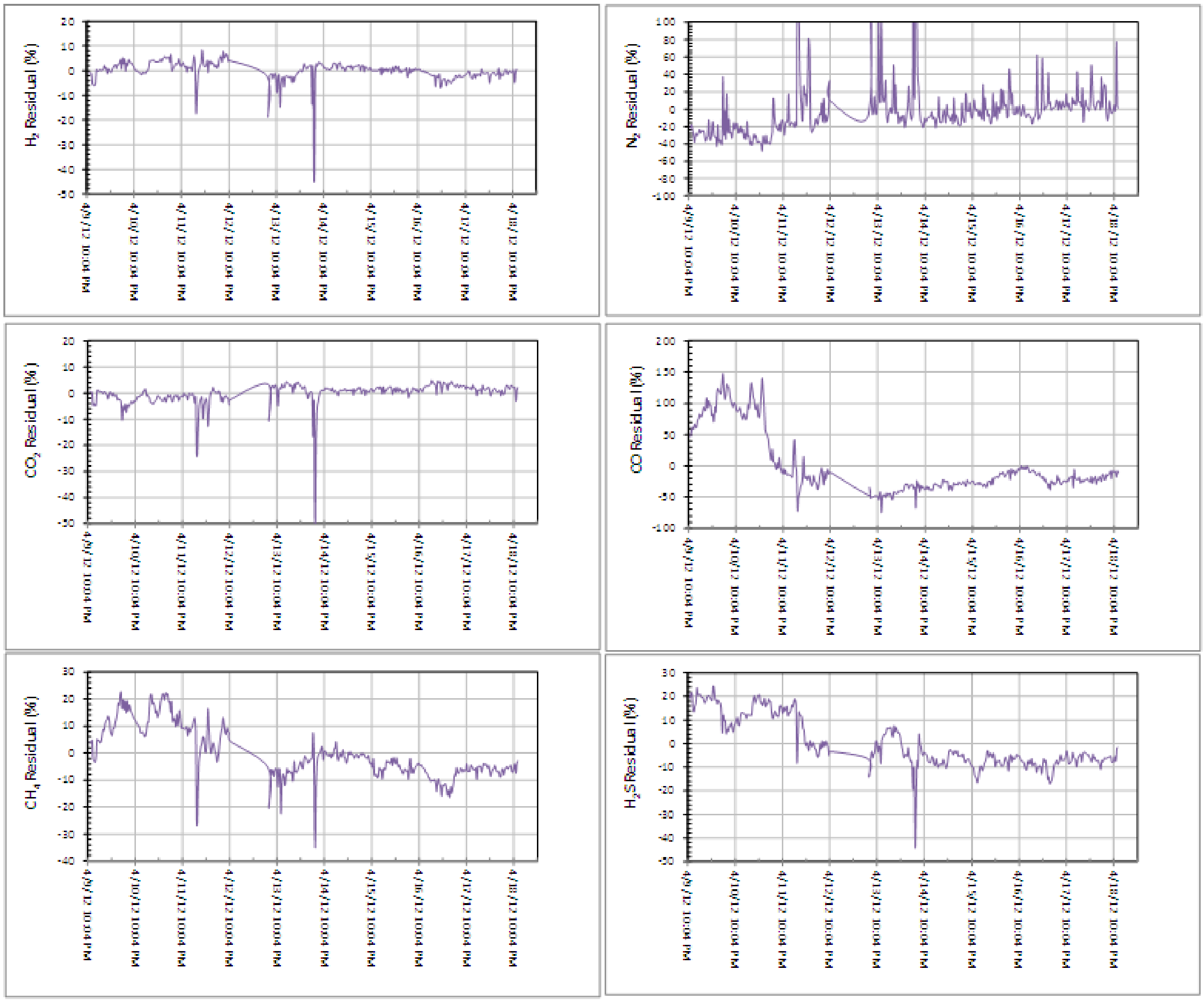


Figure T3-3. Normalized residuals for the composition of feed gas components during Campaign 2.

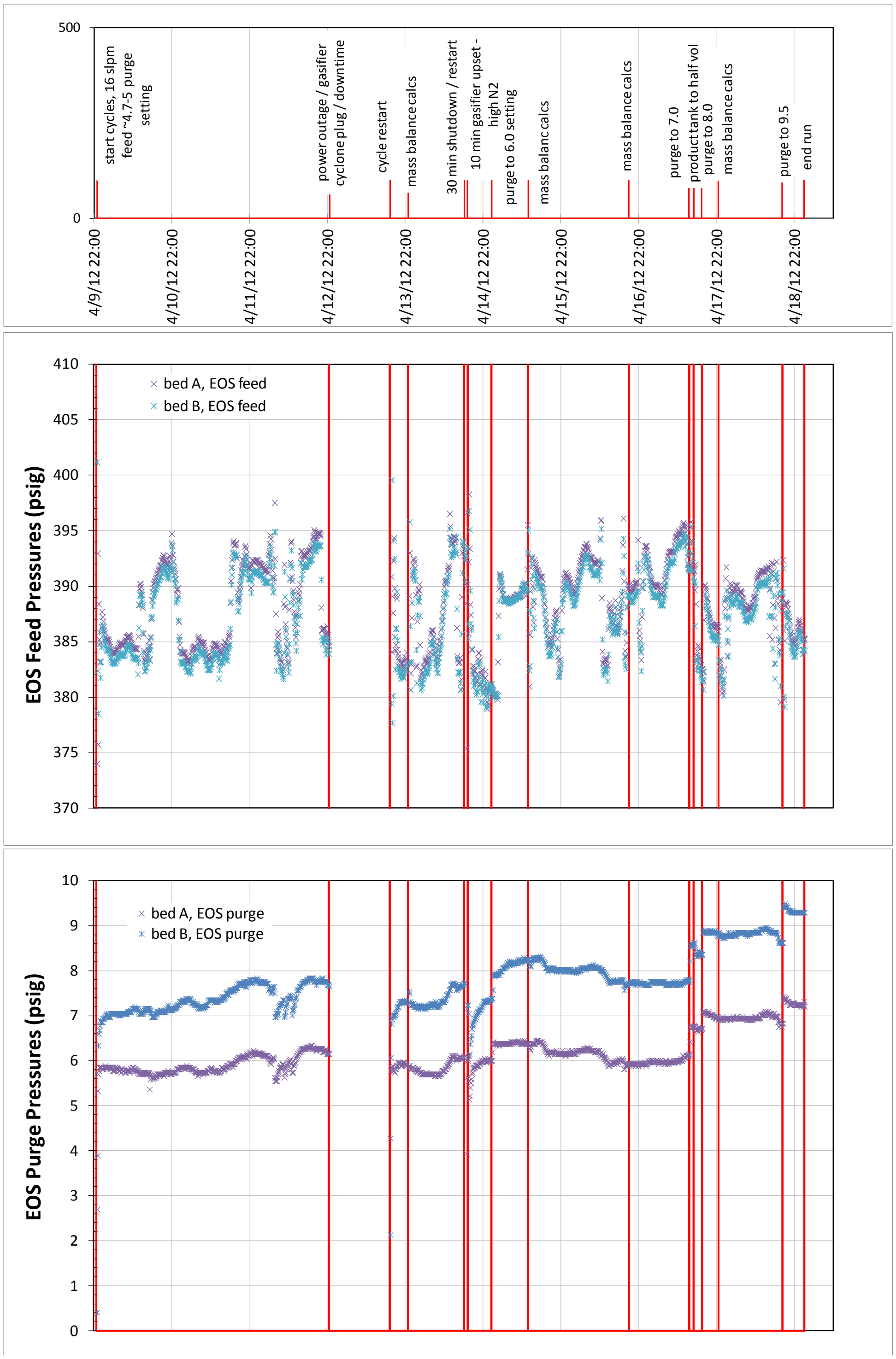


**Figure T3-4. Column pressure at the end of the feed and end of the purge step, Campaign 2.**

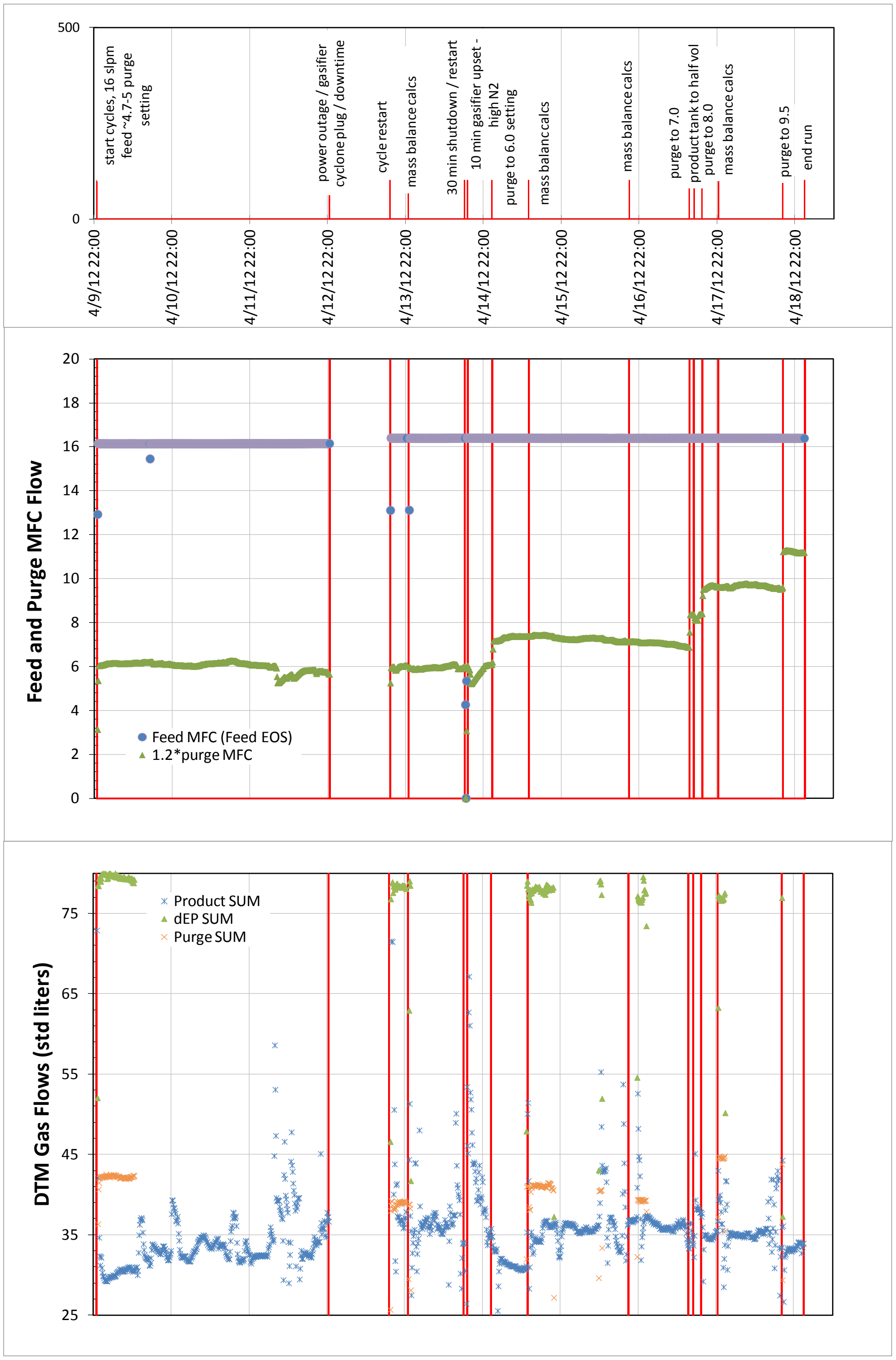
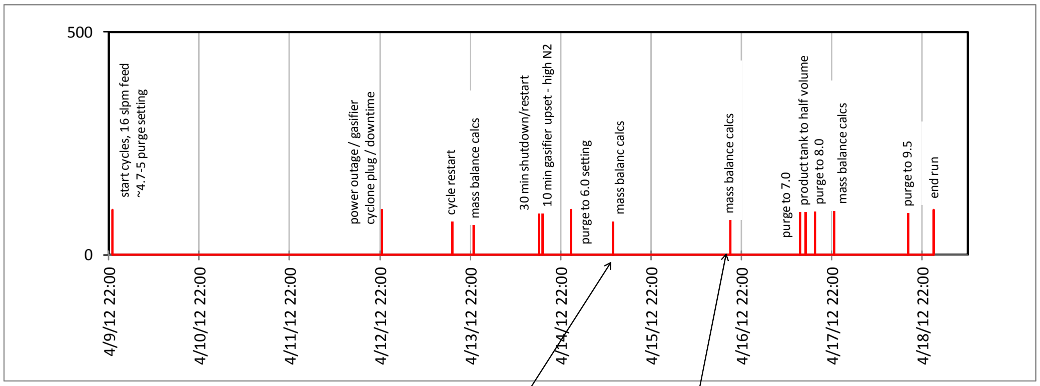


Figure T3-5. Feed and purge flow (top) and effluent flows (bottom) during Campaign 2.



| MB, In/Out | | H2 recovery | 59.91 % |
|---|---|---|---|
| total | 1.00 | H2S rejection | 99.95 % |
| H2 | 0.98 | CO2 rejection | 96.71 % |
| N2 | 0.97 | C rejection | 92.07 % |
| CO | 1.18 | 5 ppm H2S in product | |
| CO2 | 1.01 | | |
| C1 | 0.96 | | |
| C2 | 0.96 | | |
| H2S | 0.99 | | |

**jeffrey r hufton:**
missing CO in purge amount

| MB, In/Out | | H2 recovery | 58.79 % |
|---|---|---|---|
| total | 1.00 | H2S rejection | 99.95 % |
| H2 | 1.02 | CO2 rejection | 95.59 % |
| N2 | 1.12 | C rejection | 93.84 % |
| CO | 1.13 | 7 ppm H2S in product | |
| CO2 | 0.97 | | |
| C1 | 1.09 | | |
| C2 | 0.94 | | |
| H2S | 0.95 | | |

Figure T3-6. Mass Balance Evaluations during Campaign 2.

The product gas compositions during this test are illustrated in Figure T3-7. The product gas consists of about 70% $H_2$, 13% $N_2$, 2.7% CO, 8% $CO_2$, 7% $CH_4$, 0.03% $C_2H_6$. Although it is difficult to see on the log plot, Figure T3-7 shows that the $H_2$ level continuously decreases as the $CO_2$ level increases during the run.

The increasing trend in $CO_2$, CO and $H_2S$ level is more clearly illustrated in Figures T3-8, -9, and -10. $CO_2$ continuously increases at a rate of about 1 percentage point in 16 hrs. CO initially increases and declines, mimicking the trend of the feed CO concentration, and then steadily climbs at 1 percentage point in 74 hrs. These rates are essentially the same as observed at the end of Campaign 1. Neither CO nor $CO_2$ have reached a cyclic steady state in these runs.

A possible explanation for this inability to achieve CSS is associated with insufficient purge gas flow. It is important that enough purge gas is fed to the PSA to effectively regenerate the beds. This is typically quantified via a purge to feed ratio, defined as the total volume of purge gas divided by the total volume of feed gas. Conventional PSA wisdom suggests a minimum P/F ratio of 1.2 is adequate. Our P/F ratio exceeds this guideline throughout Campaign 3, indicating that the trends seen in the data cannot be from inadequate purging of the beds.

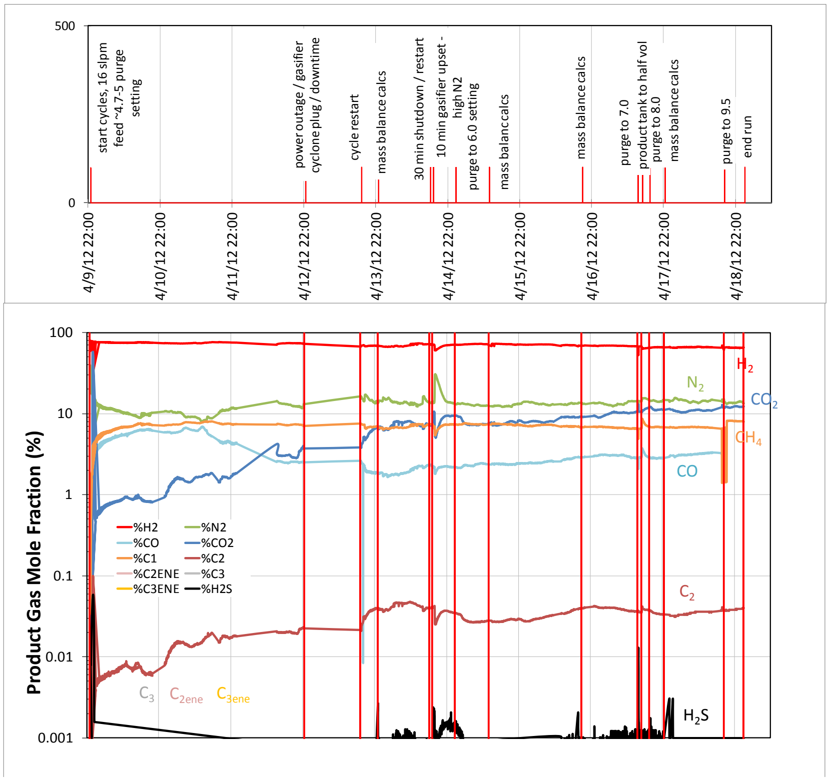


Figure T3-7. Product Gas Mole Fraction during Campaign 2.

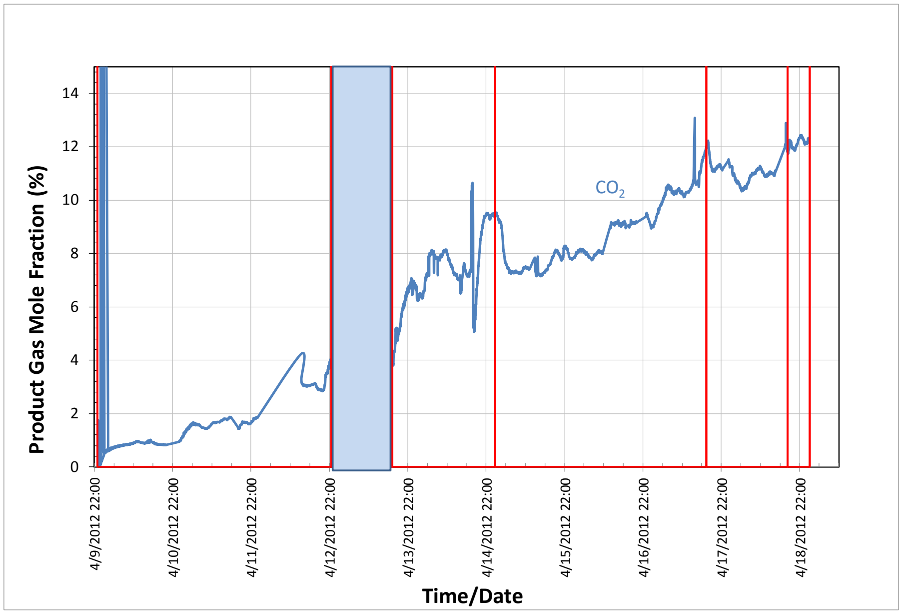


Figure T3-8. $CO_2$ Composition in the PSA Product Gas during Campaign 2.

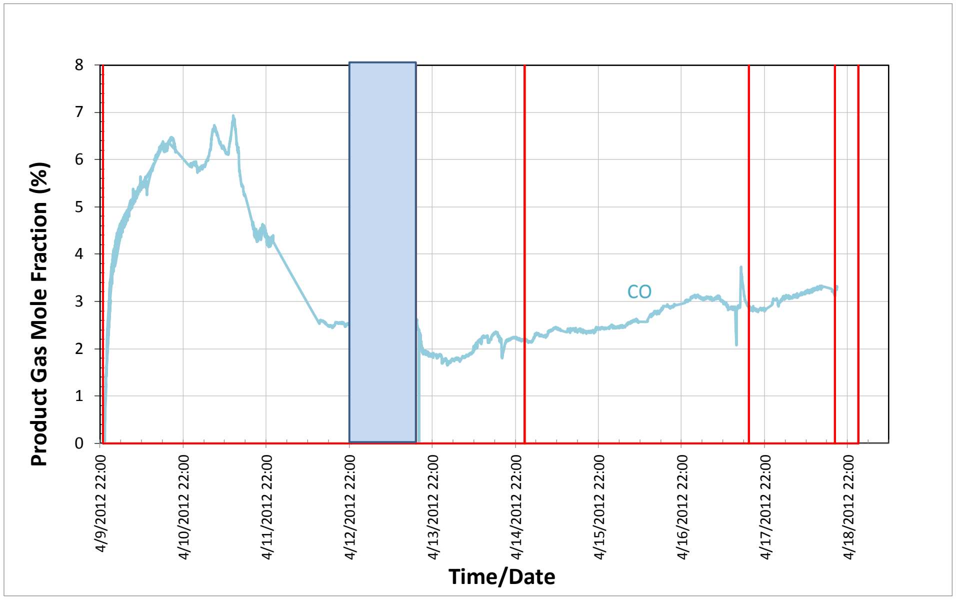


Figure T3-9. CO Composition in the PSA Product Gas during Campaign 2.

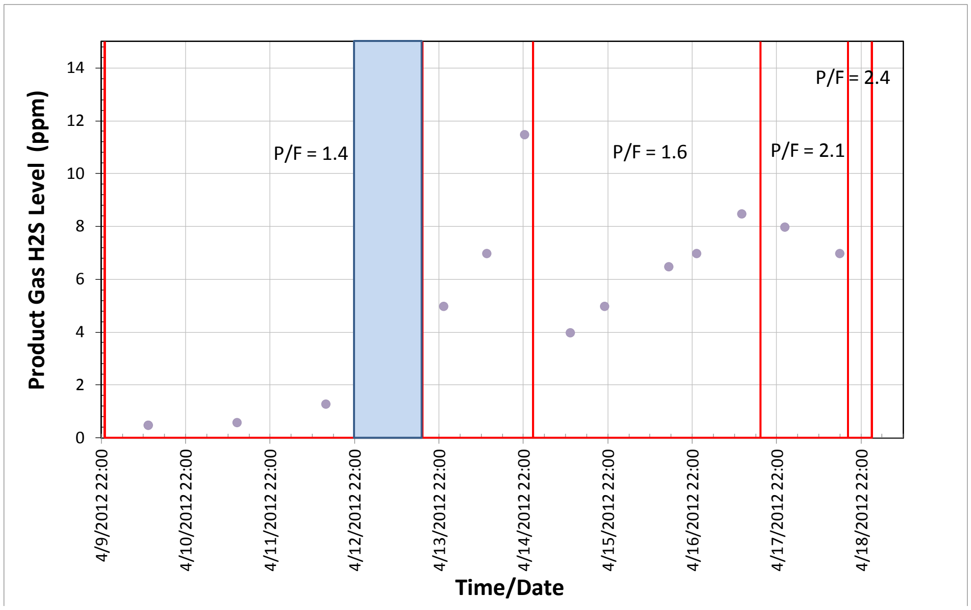


Figure T3-10. $H_2S$ Composition in the PSA Product Gas during Campaign 2.

Figure T3-11 shows the internal temperature of Bed A at the end of the feed step and at the end of the purge step. The bed heats up during the feed step as heavy species are adsorbed, and it cools when they desorb during regeneration. At cyclic steady state, the end of step (EOS) temperatures should remain constant, and the temperature difference between the feed and purge step should be constant. In the current case, the purge EOS temperature continues to increase, implying that there is less desorption of adsorbed species during the cycle. This suggests that cyclic steady state is not achieved because of difficulty in effectively regenerating the adsorbent bed.

Another proposed reason for failing to achieve CSS is based on the accumulation of tar species on the adsorbent, which eventually interfere with the adsorption of CO, $CO_2$, and $H_2S$. Some effort was directed towards the measurement of these tar species in the PSA columns. The tar GC was used to evaluate the level of organic components in the feed gas and effluent streams. It was also possible to measure the level of organics at various axial positions along the bed via taps positioned at 14, 28, 46, 64, and 82% of the bed length from the feed end. Sampling was timed so GC injection occurred at the end of the feed step, when organic levels would be highest.

Feed gas analysis via the tar GC is illustrated in Figure T3-12. Some expected components were identified and calibrated before the tests (benzene, toluene, ethyl benzene). Benzene is by far the most plentiful organic tar species with an average level of about 400 ppm.

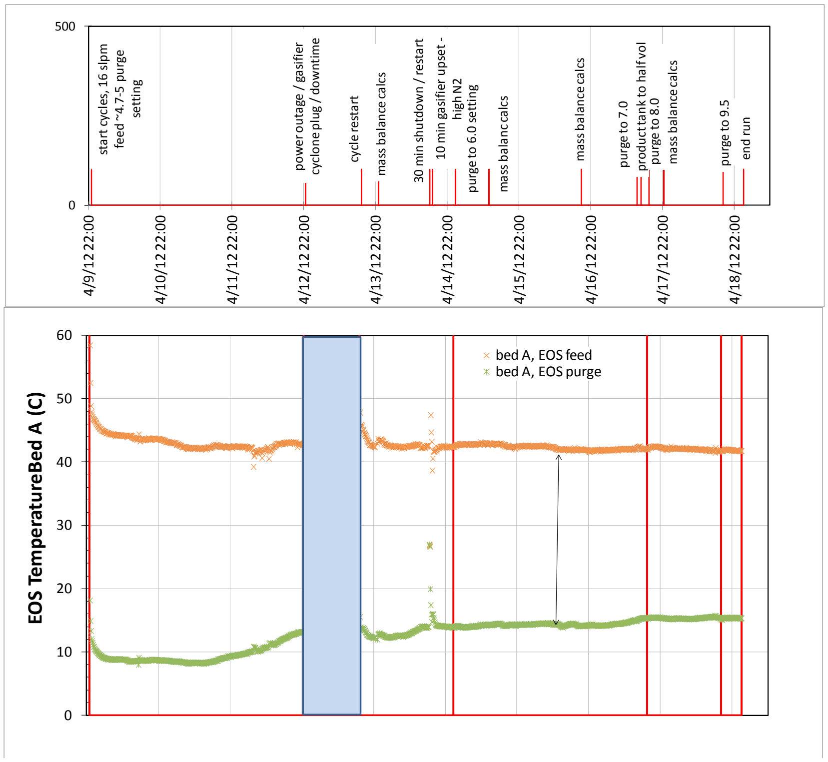
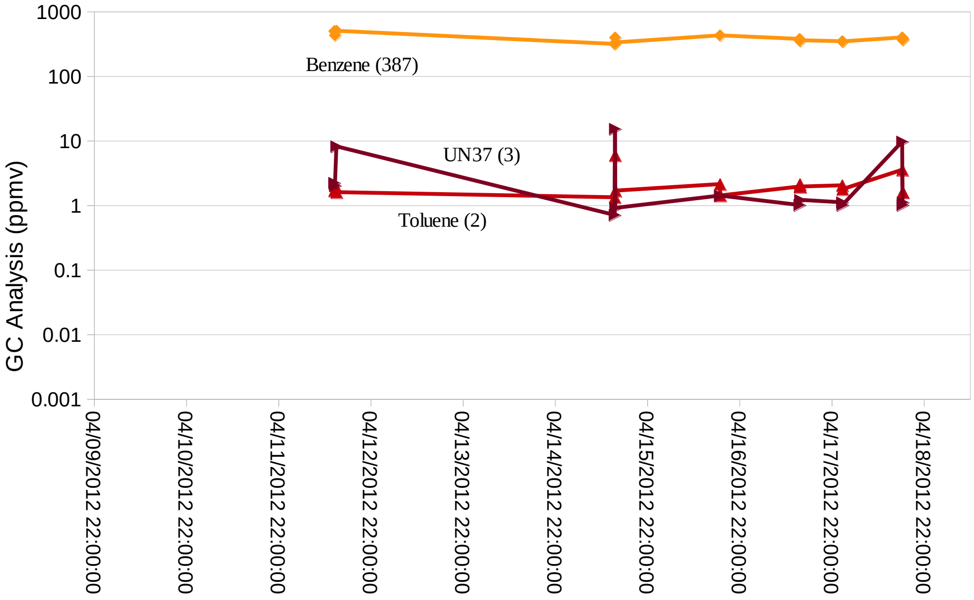


Figure T3-11. End of Step Temperature for Feed and Purge steps during Campaign 2.

Other organic components are near the limits of detection for the tar GC. Toluene is present at about 2 ppm, along with a low level of an unknown peak with a retention time of 37 seconds (UN37). Assuming the later has the same response factor as benzene, the amount of UN37 species is 3 ppm. It is most likely a substituted benzene compound.

Since significant benzene was present in the feed gas, it makes sense to look for benzene in the product gas. Analysis initially showed no organic components (i.e., below detection limit of tar GC), but during the last few days of the test breakthrough of low levels of benzene (<20 ppm) was observed (Figure T3-13). This suggests that benzene and presumably other tar species were being loaded onto the adsorbent throughout the run, and were likely responsible for the difficulty in achieving CSS.

Analysis of the gas from the various column sample taps at the end of the feed step provides some insight into the propagation of benzene during the test. Figure T3-14 shows the composition of $H_2S$ along the packed bed at the end of the feed step at various times throughout the test period. There is relatively little change in the compositions with time. Figure T3-15 illustrates the same data for benzene, where there is clear indication that the profile is moving towards the product end of the bed (right side of the figure) with continued on-stream time. This implies that the amount of benzene desorbing during regeneration is less than the amount of benzene fed, so it accumulates in the adsorbent and causes the benzene profile to travel down the column. Benzene adsorption likely interferes with $CO_2$ adsorption. The $CO_2$ profile data plotted in Figure T3-16 support this idea as $CO_2$ moves toward the product end of the bed as well. It is concluded that the presence of the organic species such as benzene (and likely many other less plentiful organic species) cannot be effectively regenerated under the current PSA cycle, and therefore accumulate on the adsorbent and interfere with adsorption of other syngas components such as $CO_2$ and CO.



**T3-12. Feed gas analysis by the tar GC during Campaign 2.**

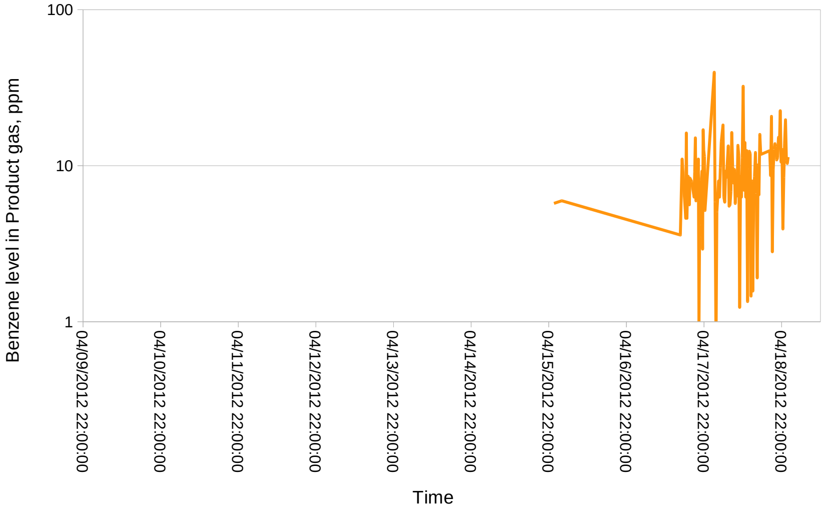


Figure T3-13. Benzene concentration in the product gas during Campaign 2.

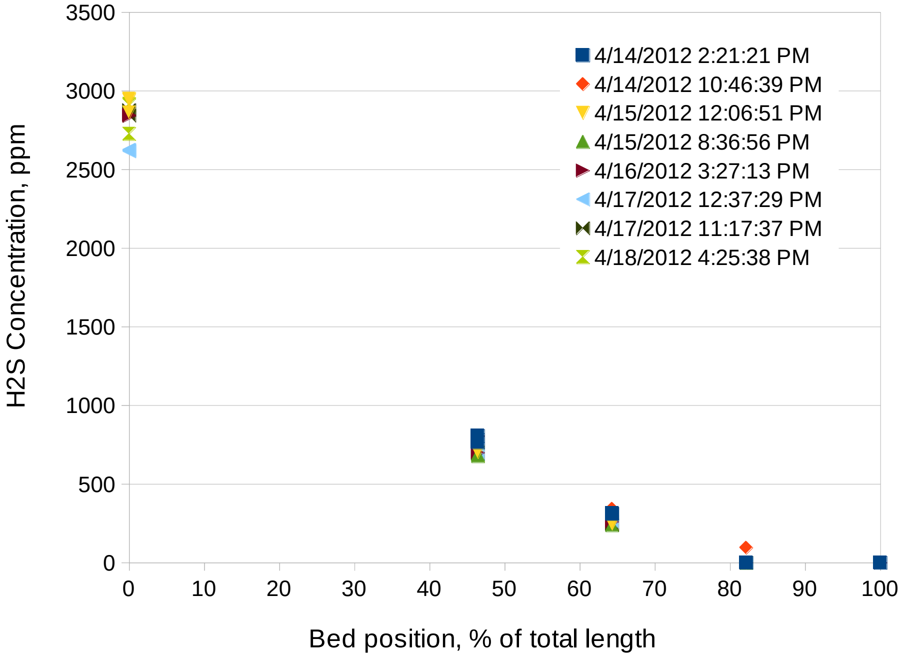


Figure T3-14. $H_2S$ Mole Fraction along the Adsorber Bed at the end of the Feed Step.

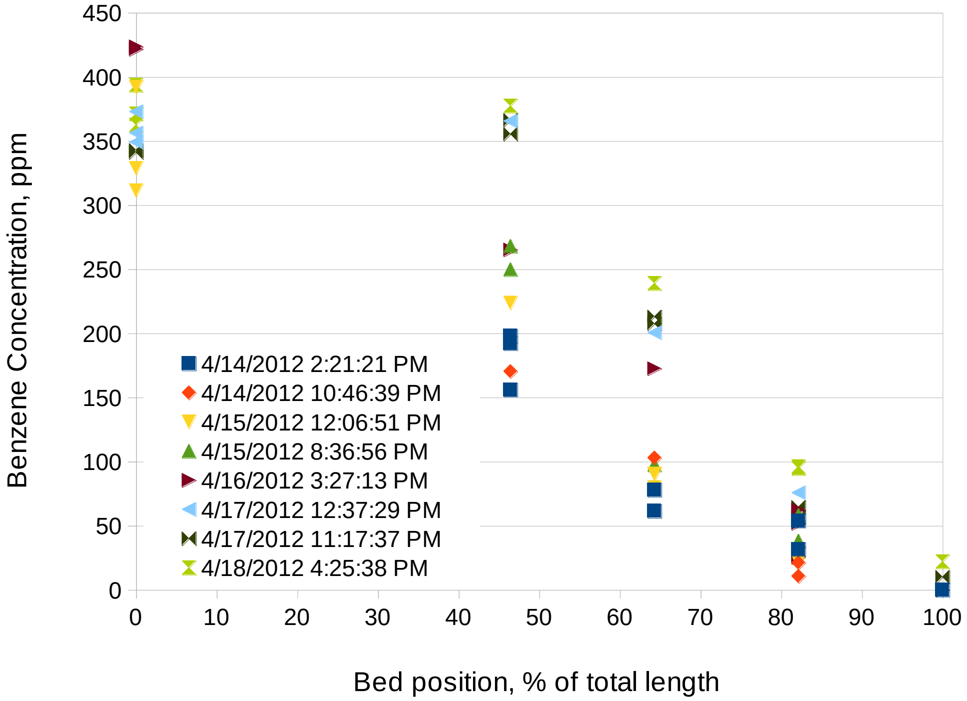


Figure T3-15. Benzene Mole Fraction along the Adsorber Bed at the end of the Feed Step.

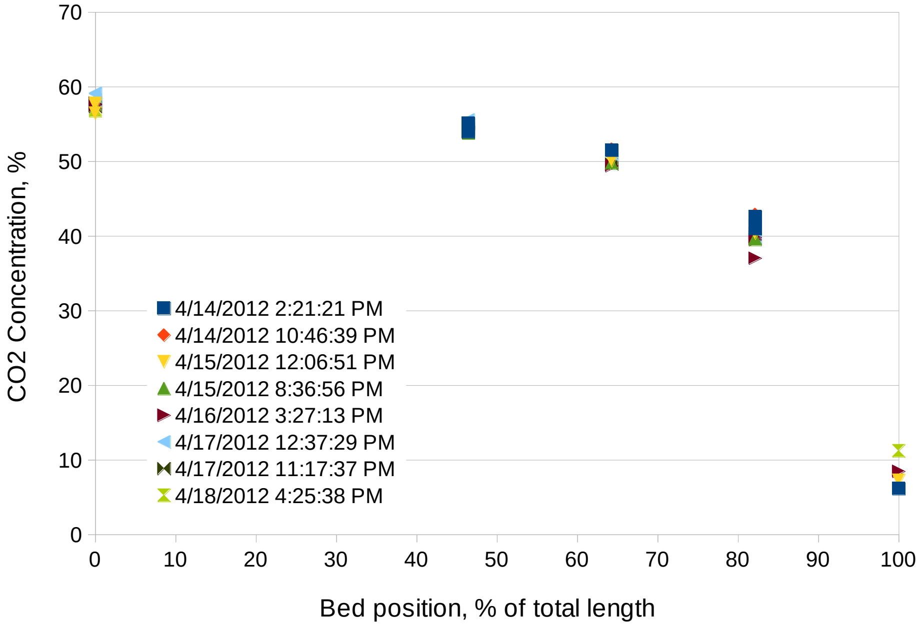


Figure T3-16. $CO_2$ Mole Fraction along the Adsorber Bed at the end of the Feed Step.

## 3- CO and $H_2S$ Breakthrough tests

CO and $H_2S$ breakthrough tests were conducted throughout the PSA testing period to gauge the impact of cycling on the adsorption capability of the adsorbent. All breakthrough tests were conducted with 10 slpm feed gas at 30°C and 400 psig, and they were conducted after the PSA vessels had been purged with $N_2$ for at least 5 hours. $H_2S$ runs were made with 2.3% $H_2S$ in $H_2$ and CO tests were conducted with 500 ppm CO in He. These tests were conducted with the fresh adsorbent, as well as after the exposures associated with Campaign 1 and 2 (see Table T3-4 for timing).

Table T3-4. Breakthrough Run Dates.

| date | BedA | BedB |
|---|---|---|
| | fresh loading | |
| **28-Feb-12** | CO | CO |
| **5-Mar-12** | H2S | H2S |
| | | |
| | campaign 1 | |
| **9-Apr-12** | | H2S |
| | campaign 2 | |
| **19-Apr-12** | H2S | |
| | thermal regen at 150C | |
| **20-Apr-12** | H2S | H2S |
| **20-May** | CO | CO |
| | remove columns | |

CO and $H_2S$ capacities are plotted in Figure T3-17. The $H_2S$ capacities are scattered by 10%, but they do not show any consistent trend after exposure to the syngas. The capacities measured after 150°C regeneration also show the same scatter. CO capacities measured after the 150°C regeneration step are essentially identical with the fresh adsorbent. It is not possible to know how the CO capacity is affected when the bed is loaded with organics as the CO detector malfunctioned during those tests. The major conclusions are 1) $H_2S$ is not affected much by exposure of the adsorbent to the raw sour syngas, and 2) after 150°C regeneration, the CO capacity of the adsorbent is retained. There appears to be no adsorbent degradation or pore blocking associated with adsorbent contact with the sour syngas.

## 4- PSA Adsorbent Post-Mortem Analysis

At the end of Campaign 2, the two PSA vessels were thermally regenerated under flowing $N_2$ at 150°C (described later), sealed, and sent to APCI. They were carefully unloaded in a ventilated hood. Adsorbent samples were removed in increments from the feed end of the bed and are listed in Table T3-5. Entries in red text were submitted for various lab analyses. Samples from both vessels exhibited a color gradient from the feed (dark) to product (colorless) end of the vessels (Figure T3-18) .

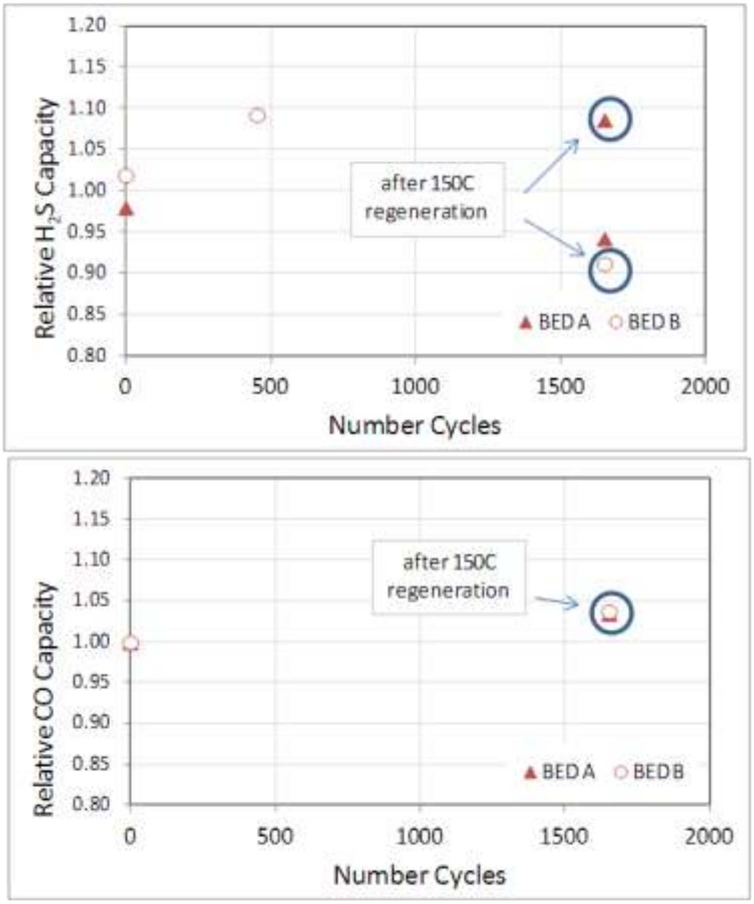


Figure T3-17. Relative $H_2S$ and CO Capacities Determined by Breakthrough Experiments During Campaigns 1 and 2.

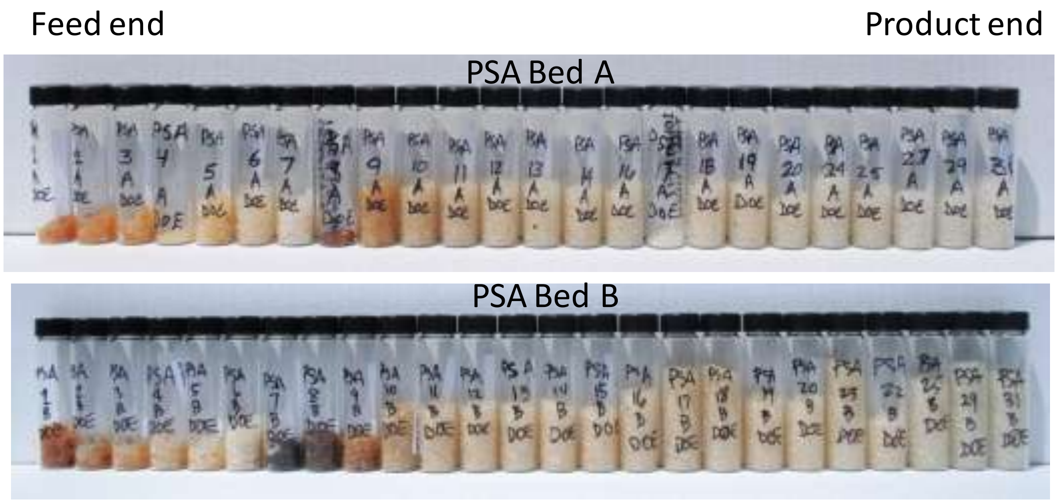


Figure T3-18. Photo of adsorbent samples removed from the PSA columns after Campaign 2.

Changes in the adsorption properties of the samples were evaluated via TGA experiments. Weight losses upon heating the samples to 200°C in $N_2$ and $CO_2$ capacities at 40°C are listed in Table T3-6. The thermal regeneration at EERC appears to have eliminated most of the accumulated organics, since there is relatively little weight loss on heating to 200°C. Evolution of gases during heating to 200°C was monitored by mass spectrometry (MS) and only water and a rare trace of $CO_2$ were observed. No hydrocarbons were detected.

There was almost no change in $CO_2$ capacities for the ADS_C samples and only modest decline for the ADS_A Bed A samples. Capacities decreased by about 8 to 13% for the ADS_A samples from Bed B. For the ADS_C, a maximum capacity loss of about 10% was obtained for the sample at the outlet end of the ADS_C layer.

XRF data was obtained for selected samples from Bed A (Table T3-7). Sulfur concentrations were only slightly greater than those of the fresh adsorbents, but there appears to be a slight sulfur gradient. There was no significant difference in concentrations versus the fresh adsorbents for any other element.

## Table T3-5. Adsorbent Samples from PSA Bed A and Bed B.

| Bed A DOE #2 | | | Bed B DOE #2 | | |
|---|---|---|---|---|---|
| **sample #** | **Adsorbent** | **depth, in.** | **sample #** | **adsorbent** | **depth, in.** |
| **20120724-1** | ADS_C | | **20120725-1** | ADS_C | |
| **20120724-2** | ADS_C | | **20120725-2** | ADS_C | |
| **20120724-3** | ADS_C | | **20120725-3** | ADS_C | |
| **20120724-4** | ADS_C | 2 | **20120725-4** | ADS_C | 2.5 |
| **20120724-5** | ADS_C | | **20120725-5** | ADS_C | |
| **20120724-6** | ADS_C | | **20120725-6** | ADS_C | |
| **20120724-7** | ADS_C | 5 | **20120725-7** | ADS_A | |
| **20120724-8** | ADS_A | | **20120725-8** | ADS_A | |
| **20120724-9** | ADS_A | | **20120725-9** | ADS_A | |
| **20120724-10** | ADS_A | 9.5 | **20120725-10** | ADS_A | 9 |
| **20120724-11** | ADS_A | | **20120725-11** | ADS_A | |
| **20120724-12** | ADS_A | | **20120725-12** | ADS_A | |
| **20120724-13** | ADS_A | | **20120725-13** | ADS_A | |
| **20120724-14** | ADS_A | | **20120725-14** | ADS_A | |
| **20120724-15** | ADS_A | 16.5 | **20120725-15** | ADS_A | 14.5 |
| **20120724-16** | ADS_A | | **20120725-16** | ADS_A | |
| **20120724-17** | ADS_A | | **20120725-17** | ADS_A | |
| **20120724-18** | ADS_A | | **20120725-18** | ADS_A | |
| **20120724-19** | ADS_A | | **20120725-19** | ADS_A | |
| **20120724-20** | ADS_A | 23 | **20120725-20** | ADS_A | 28 |
| **20120724-21** | ADS_A | | **20120725-21** | ADS_A | |
| **20120724-22** | ADS_A | 33 | **20120725-22** | ADS_A | |
| **20120724-23** | ADS_A | | **20120725-23** | ADS_A | 41.5 |
| **20120724-24** | ADS_A | 46.5 | **20120725-24** | ADS_A | 54 |
| **20120724-25** | ADS_A | | **20120725-25** | ADS_A | |
| **20120724-26** | ADS_A | 66 | **20120725-26** | ADS_A | no sample |
| **20120724-27** | ADS_A | | **20120725-27** | ADS_A | no sample |
| **20120724-28** | ADS_A | 74 | **20120725-28** | ADS_A | * |
| **20120724-29** | ADS_A | | **20120725-29** | ADS_A | * |
| **20120724-30** | ADS_A | | **20120725-30** | ADS_A | * |
| **20120724-31** | ADS_A | | **20120725-31** | ADS_A | * |
| **20120724-32** | ADS_A | | **20120725-32** | ADS_A | * |
| *** hose stuck - sample removed from exit end** | | | | | |

### Table T3-6. TGA results for adsorbent samples from PSA Bed A and B. 200°C $N_2$, 30 min; 40°C, 30 min 100% $CO_2$; 30 min $N_2$.

| Bed A samples | adsorbent | wt% loss 200°C $N_2$ | mmol$CO_2$/g adsorbed | % change vs fresh | mmol $N_2$/g 40°C[a] | % change vs. fresh |
|---|---|---|---|---|---|---|
| **20120724-1** | ADS_C | 1.3156 | 0.266 | 16.9 | 0.0199 | 8.9 |
| **20120724-2** | ADS_C | 2.3006 | 0.242 | 6.2 | 0.0182 | 0.0 |
| **20120724-4** | ADS_C | 3.0007 | 0.255 | 11.7 | 0.0204 | 11.6 |
| **20120724-7** | ADS_C | 1.3016 | 0.207 | -9.2 | 0.0182 | -0.3 |
| **20120724-8** | ADS_A | 0.5797 | 0.790 | -2.3 | 0.0436 | 17.1 |
| **20120724-9** | ADS_A | 0.6045 | 0.751 | -7.2 | 0.0415 | 11.6 |
| **20120724-10** | ADS_A | 0.6130 | 0.750 | -7.3 | 0.0437 | 17.4 |
| **20120724-16** | ADS_A | 1.1204 | 0.760 | -6.1 | 0.0429 | 15.4 |
| **20120724-22** | ADS_A | 1.0186 | 0.723 | -10.6 | 0.0402 | 8.1 |
| **20120724-28** | ADS_A | 1.0424 | 0.763 | -5.7 | 0.0402 | 7.9 |
| **Bed B samples** | **adsorbent** | **wt% loss 200°C $N_2$** | **mmol$CO_2$/g adsorbed** | **% change vs fresh** | **mmol $N_2$/g 40°C[a]** | **% change vs. fresh** |
| **20120725-1** | ADS_C | 1.4961 | 0.232 | 1.8 | 0.0177 | -2.8 |
| **20120725-2** | ADS_C | 0.7761 | 0.219 | -3.9 | 0.0158 | -13.6 |
| **20120725-6** | ADS_C | 1.0478 | 0.206 | -9.7 | 0.0169 | -7.4 |
| **20120725-7** | ADS_A | 1.0198 | 0.703 | -13.1 | 0.0409 | 9.8 |
| **20120725-8** | ADS_A | 0.7915 | 0.717 | -11.4 | 0.0416 | 11.7 |
| **20120725-9** | ADS_A | 0.7783 | 0.744 | -8.0 | 0.0420 | 12.9 |
| **20120725-32** | ADS_A | 0.7004 | 0.717 | -11.4 | 0.0454 | 22.0 |

**a. $N_2$ uptake after cooling from 200 to 40°C.**

### Table T3-7. XRF analysis for selected Bed A samples.

| sample | adsorbent | wt% S | mmol S/g |
|---|---|---|---|
| **fresh** | ADS_C | 0.001 | 0.0003 |
| **20120724-1** | ADS_C | 0.0168 | 0.0052 |
| **20120724-7** | ADS_C | 0.0116 | 0.0036 |
| **fresh** | ADS_A | 0.0329 | 0.0103 |
| **20120724-8** | ADS_A | 0.0921 | 0.0287 |
| **20120724-10** | ADS_A | 0.0641 | 0.0200 |

## 5- Desorption of Tar Species from PSA Adsorbents

Once it was shown that tar species interfere with the operation of the sour PSA system, it became important to understand 1) if and how the tar-loaded adsorbent could be regenerated and 2) whether an additional separation system could be used to remove the tars before the syngas was sent to the sour PSA. Likewise, it was instructive to determine which tar species were present, and what adsorbents were effective in removing them.

The first question was addressed with the sour PSA unit at the end of Campaign 2. At the end of the PSA cycling, the two beds were regenerated by first purging with 20 slpm $N_2$ (~10 slpm each bed) at atmospheric pressure and 30°C for 9 hours. They were then externally heated to 150°C, held at that temperature for 15 hrs, and cooled back to 30°C (total of 20 hrs), all under the same $N_2$ purge flow rate. The concentration of organics and heavier syngas components in the effluent stream was monitored, and the results are plotted in Figure T3-19 (purge) and Figure T3-20 (heating). About 40% of the benzene comes out during the 30°C purge. It is reversibly adsorbed, but purging at 30°C is not very efficient. Excessive time and $N_2$ would be needed to remove the benzene – higher molecular weight species would require even more time. Benzene, toluene, and UN37 are still present in the bed at the end of the 30°C purge step. The syngas components $CO_2$, $H_2S$, and ethane readily desorb via the simple $N_2$ purge step.

Desorption of the organics is enhanced by heating the bed. Benzene levels reach a maximum concentration of 4500 ppm (still substantially below the dew point concentration at room temperature of around 3%). Other adsorbed organics become evident during the heating step – e.g. xylene, and two unidentified species UN70 and UN185. They were too dilute to see in the feed gas, but the concentrating effect of the adsorption

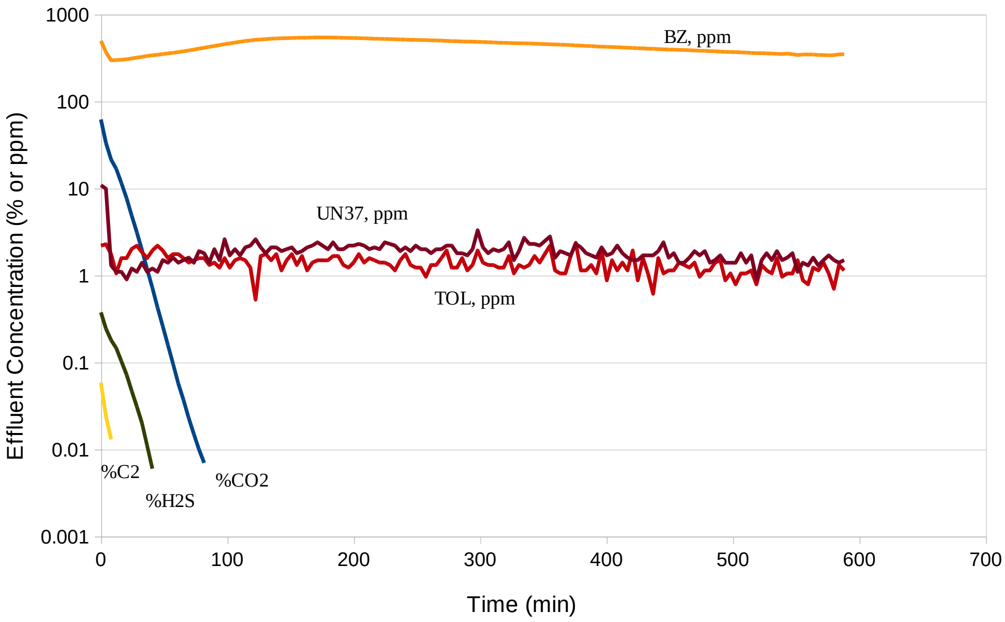


Figure T3-19. Effluent Composition from PSA beds after Campaign 2 purged with $N_2$ at 30°C.

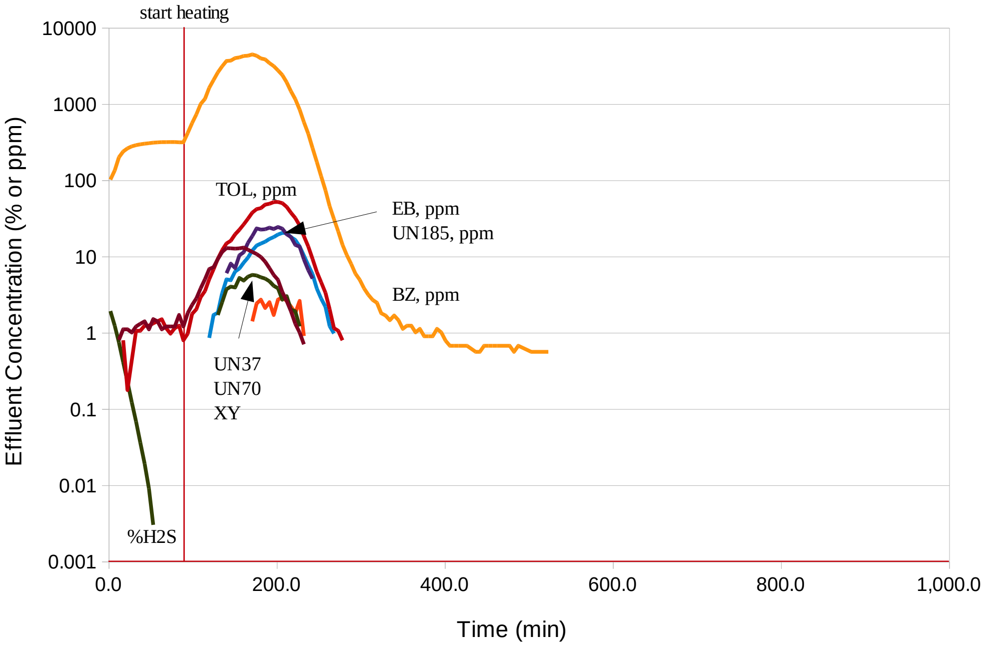


**Figure T3-20. Effluent Composition from PSA beds after Campaign 2 purged with $N_2$ and heated to 150°C.**

bed enables their detection in the effluent gas. Most of the benzene is gone after about 350 min (~6 hrs) of heating. This test indicates that thermal regeneration of the PSA adsorbent is capable of effectively desorbing the tar components.

### 6- Behavior of Tar Species on Various Adsorbents

Thermal regeneration can be used as an approach to handle PSA adsorbent that has become contaminated with tar species during PSA operation, or it could be the operating approach of a separate upstream adsorption system designed to remove the tar components before the sour PSA. In this case it is useful to understand adsorption of tars on other adsorbents, as there may be better adsorbents than ADS_A and ADS_B for removing the large organics.

The six-bed test apparatus was used to evaluate exposure of adsorbents to the sour tar-laden syngas. The idea was to completely saturate the adsorbent samples with tar-containing syngas, then remove the samples and analyze them via various lab techniques. Each bed contained 0.5-1.5 g sample of the adsorbent of interest at the inlet end, followed by a 0.5 g sample of ADS_F and a layer of 0.5 mm glass beads (Figure T3-21). The ADS_F layer was included to asses if uniform gas flow was maintained in each of the parallel packed columns. An exposure run was completed during each campaign: Exposure 1 conducted during Campaign 1 with adsorbent exposure for 77 hrs; Exposure 2 conducted during Campaign 2 with adsorbent exposure for 150 hrs; and Exposure 3 conducted during Campaign 3 with adsorbent exposure for 82 hrs. The adsorbent samples were removed from the test apparatus and analyzed after each exposure, and fresh samples were tested in each run. The adsorbents tested and their initial activation conditions are listed in Table T3-8.

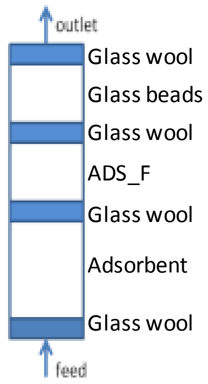


Figure T3-21. Arrangement of materials in the six-bed test apparatus.

After syngas exposure, samples were unloaded in a $N_2$ purged dry box. Selected samples from Exposures 2 and 3 were analyzed by XRF to determine elemental composition. Relative to the fresh adsorbents, the only element that changed in concentration was sulfur (Table T3-9). Adsorbents ADS_E and ADS_D had the highest sulfur loadings while those on the PSA adsorbents ADS_A and ADS_B were the lowest.

TGA testing was conducted with the loaded adsorbent samples as well as the ADS_F traps. $CO_2$ capacities and weight losses after heating to 200°C are listed in Table T3-10. The weight losses (percentage basis) for the ADS_F traps were about the same for each run, supporting equal syngas exposure for each bed. Decreases in $CO_2$ capacities relative to those of the fresh adsorbents were, in general, modest. With the exception of ADS_C, weight losses for the adsorbent of interest were rather large, typically 10% or more. Mass spectroscopy analysis of the effluent gas from the TGA indicated the absence of $NH_3$, HCl, $H_2S$, and COS while $CO_2$ and $H_2O$ were detected for some samples. In addition, organic species were identified including benzene, toluene, indane, naphthalene, and tetrahydronaphthalene, as well as a few mass peaks of unknown origin.

To obtain a more thorough characterization of adsorbed species, samples from Exposure 3 were analyzed by GCMS. The adsorbents were heated to 150°C for 30 min and 2.5 mL of the headspace gases were used for analysis. ADS_G was heated to 140°C to avoid any decomposition products. Table T3-11 provides a summary of the species detected. As indicated by the molecular structures at the bottom of that table, some of the observed species are very large aromatic or polyaromatic species that would be expected to be strongly adsorbed on the various adsorbents.

## Table T3-8. Adsorbents and activation conditions.

| Adsorbents | Activation |
|---|---|
| **ADS_A, 0.5-1 mm beads, 63231-67-4** | 150°C, $N_2$ purge overnight |
| **ADS_H, 1 mm beads** | 150°C, $N_2$ purge overnight |
| **ADS_C, crushed and sieved 20-16 mesh** | 150°C, $N_2$ purge overnight |
| **ADS_B, crushed and sieved 20-16 mesh** | 150°C, $N_2$ purge overnight |
| **ADS_G, small beads** | washed and dried previously, 100°C overnight |
| **ADS_E, crushed and sieved 20-16 mesh** | 200°C, $N_2$ purge 3 hours |
| **ADS_D, crushed and sieved 20-16 mesh** | None |

## Table T3-9. Sulfur concentrations (XRF) for fresh and spent adsorbents.

| | ADS_A | | ADS_B | | ADS_E | |
|---|---|---|---|---|---|---|
| **Exp 2** | **fresh** | **spent 04-5** | **fresh** | **spent 04-3** | **Fresh** | **spent 04-2** |
| **wt% S** | 0.0329 | 0.097 | 0.0307 | 0.076 | 0.749 | 1.0 |
| **mmol S/g** | 0.010 | 0.030 | 0.010 | 0.024 | 0.234 | 0.312 |
| | **ADS_D** | | **ADS_C** | | **ADS_G** | |
| **Exp 3** | **fresh** | **spent 511-2** | **fresh** | **wt% S** | **fresh** | **spent 511-2** |
| **wt% S** | 0.010 | 1.2 | 0.0010 | wt% | 0.010 | 1.2 |
| **mmol S/g** | 0.0037 | 0.37 | 0.00031 | mmol S/g | 0.0037 | 0.37 |

The next test was designed to determine how easily the adsorbed organics could be desorbed from the ADS_A and ADS_G samples from Exposure 1. The first portion of regeneration mimicked those associated with a pressure swing adsorption cycle and consisted of simple purging with $N_2$ at 40°C for 2 h. After that, the samples were heated at 10°C/min to 200°C for ADS_A and 100°C for ADS_G. Desorption occurred more readily for the ADS_G adsorbent, with desorption of almost 80% of the adsorbed gases during the 40°C $N_2$ purge versus only 64% for ADS_A (Table T3-12) . These tests suggest that 40°C was inadequate to desorb organics from adsorbent ADS_A. This is consistent with the results of the PSA testing with the sour PSA unit.

Two additional weight loss experiments were conducted to determine if ADS_A can be fully regenerated at 150°C rather than 200°C. The results in Table T3-13 show that 1 hour at 150°C was too short to achieve complete regeneration while 3 hours at 150°C was sufficient to achieve 98.1% regeneration. Thus, regeneration at 150°C may be adequate if the purge time is long enough.

## Table T3-10. TGA results for spent adsorbents and ADS_F traps.

| **Run 1** | | | | | | |
|---|---|---|---|---|---|---|
| **bed #** | **sample** | **adsorbent** | **wt% loss 200°C $N_2$** | **mmol$CO_2$/g adsorbed** | **% change vs fresh** | **mmol $N_2$/g 40°C[a]** |
| **1** | 20120301-1 | ADS_A | 14.787 | 0.678 | -16.2 | 0.0716 |
| **2** | 20120301-2 | ADS_C | 3.429 | 0.237 | 4.14 | 0.0128 |
| **3** | 20120301-3 | ADS_B | 12.962 | 0.686 | 0.29 | 0.0339 |
| **4** | 20120301-4 | ADS_H | 9.379 | 0.776 | 6.77 | 0.0246 |
| **5** | 20120301-5 | ADS_G | 18.692 | 0.648 | -18.0 | 0.0241 |
| **6** | 20120301-6 | ADS_E | 19.285 | 0.990 | -0.3 | 0.159 |
| **1** | 20120301-1C | ADS_F | 20.300 | 1.409 | -14.1 | 0.216 |
| **2** | 20120301-2C | ADS_F | 21.469 | 1.339 | -18.4 | 0.211 |
| **3** | 20120301-3C | ADS_F | 24.165 | 1.595 | -2.8 | 0.281 |
| **4** | 20120301-4C | ADS_F | 21.514 | 1.473 | -10.2 | 0.242 |
| **5** | 20120301-5C | ADS_F | 20.280 | 1.524 | -7.1 | 0.245 |
| **6** | 20120301-6C | ADS_F | 20.717 | 1.550 | -5.5 | 0.248 |
| **Run 2** | | | | | | |
| **bed #** | **sample** | **adsorbent** | **wt% loss 200°C $N_2$** | **mmol$CO_2$/g adsorbed** | **% change vs fresh** | **mmol $N_2$/g 40°C[a]** |
| **5** | 20120404-5 | ADS_A | 16.562 | 0.676 | -16.4 | 0.0375 |
| **2** | 20120404-2 | ADS_E | 14.353 | 0.598 | -39.8 | 0.0800 |
| **3** | 20120404-3 | ADS_B | 15.409 | 0.669 | -2.1 | 0.0347 |
| **4** | 20120404-4 | ADS_D | 26.677 | 0.843 | -22.4 | 0.1122 |
| **none** | fresh | ADS_D | 3.427 | 1.086 | - | 0.1561 |
| **5** | 20120404-5C | ADS_F | 18.5185 | 1.275 | -22.3 | 0.1895 |
| **2** | 20120404-2C | ADS_F | 21.6649 | 1.451 | -11.5 | 0.2331 |
| **3** | 20120404-3C | ADS_F | 19.7546 | 1.344 | -18.1 | 0.2082 |
| **4** | 20120404-4C | ADS_F | 20.3699 | 1.434 | -12.6 | 0.2174 |
| **Run 3** | | | | | | |
| **bed #** | **sample** | **adsorbent** | **wt% loss 200°C $N_2$** | **mmol$CO_2$/g adsorbed** | **% change vs. fresh** | **mmol $N_2$/g 40°C[a]** |
| **1** | 20120511-1 | ADS_B | 13.5588 | 0.710 | 3.9 | 0.0453 |
| **2** | 20120511-2 | ADS_D | 17.9799 | 0.890 | -18.1 | 0.0128 |
| **3** | 20120511-3 | ADS_A | 12.6369 | 0.726 | -10.3 | 0.1271 |
| **4** | 20120511-4 | ADS_C | 3.422 | 0.249 | 9.3 | 0.0236 |
| **5** | 20120511-5 | ADS_E | 15.2284 | 0.755 | -23.9 | 0.1266 |
| **6** | 20120511-6 | ADS_G | 13.5845 | 0.709 | -10.3 | 0.0322 |
| **1** | 20120511-1C | ADS_F | 21.4289 | 1.583 | -3.5 | 0.264 |
| **2** | 20120511-2C | ADS_F | 19.1059 | 1.543 | -5.9 | 0.253 |
| **3** | 20120511-3C | ADS_F | 20.0930 | 1.550 | -5.5 | 0.254 |
| **4** | 20120511-4C | ADS_F | 17.1162 | 1.138 | -30.7 | 0.171 |
| **5** | 20120511-5C | ADS_F | 19.3393 | 1.609 | -1.9 | 0.279 |
| **6** | 20120511-6C | ADS_F | 20.1230 | 1.581 | -3.6 | 0.276 |

a. $N_2$ uptake after cooling from 200 to 40°C.

### Table T3-11. Adsorbed species identified by GCMS analysis.

*** Structural formula number; structures after table.**

| | * | adsorbents containing indicated species | | | | | | |
|---|---|---|---|---|---|---|---|---|
| **Aromatics:**<br>**benzene** | | ADS_A | ADS_B | ADS_C | ADS_H | ADS_E | ADS_D | ADS_G |
| **Toluene** | | ADS_A | ADS_B | ADS_C | ADS_H | ADS_E | ADS_D | ADS_G |
| **Ethylbenzene** | | ADS_A | ADS_B | ADS_C | ADS_H | ADS_E | ADS_D | ADS_G |
| **ethylmethylbenzene** | | ADS_A | ADS_C | ADS_C | ADS_H | ADS_E | | |
| **Xylene** | | ADS_A | ADS_B | ADS_C | | ADS_E | | ADS_G |
| **Propylbenzene** | | ADS_A | | | | | | |
| **C3 benzene** | | | | | | ADS_E | | ADS_G |
| **Indane** | 8 | ADS_A | ADS_B | ADS_C | ADS_H | ADS_E | ADS_D | ADS_G |
| **Methylindane** | 9 | ADS_A | | ADS_C | ADS_H | | | |
| **Naphthalene** | 10 | ADS_A | ADS_B | ADS_C | ADS_H | ADS_E | ADS_D | ADS_G |
| **methylnaphthalene** | 11 | ADS_A | | ADS_C | | | | ADS_G |
| **1,2,3,4-tetrahydronaphthalene** | 12 | ADS_A | ADS_B | ADS_C | ADS_H | ADS_E | ADS_D | ADS_G |
| **1,2,2a,3,4,5-hexahydroacenaphthylene** | 13 | ADS_A | | ADS_C | | | | |
| **1,2 dihydroacenaphthylene** | 14 | | | ADS_C | | | | ADS_G |
| **9H-fluorene** | 15 | | | ADS_C | | | | |
| **Cyclopentene** | 21 | | | | | | | ADS_G |
| **1-1'-biphenyl** | 22 | | | | | | | ADS_G |
| **Oxygen containing:**<br>**dibenzofuran** | 16 | | | ADS_C | | | | |
| **Benzofuran** | 23 | | | | | | | ADS_G |
| **Nitrogen containing: pyridine** | 17 | | | ADS_C | ADS_H | ADS_E | ADS_D | |
| **2,4-pentadienenitrile** | 24 | | | | | | | ADS_G |
| **Sulfur containing:**<br>**carbon disulfide** | 18 | | | | | ADS_E | | |
| **Thiophene** | 19 | ADS_A | ADS_B | ADS_C | ADS_H | ADS_E | ADS_D | ADS_G |
| **benzo[b]thiophene** | 20 | ADS_A | ADS_B | ADS_C | ADS_H | ADS_E | | ADS_G |

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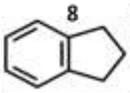

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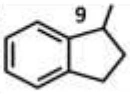

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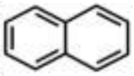

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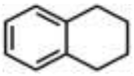

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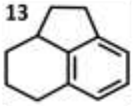

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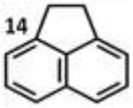

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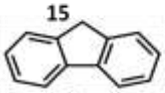

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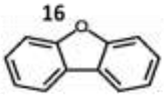

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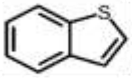

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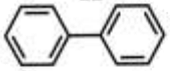

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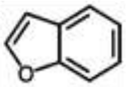

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Table T3-12. TGA weight losses for ADS_A and ADS_G under $N_2$ purge at 40°C.

| | ADS_A | ADS_G |
|---|---|---|
| wt% loss, 40°C, 2h | 9.306 | 14.438 |
| wt% loss, 40°C to 200°C or 100°C , 30 min | 5.334 (200°C) | 3.667 (100°C) |
| total wt% loss | 14.641 | 18.104 |
| % of total desorbed at 40°C | 63.6 | 79.7 |

Table T3-13. Weight losses for heating spent ADS_A to 150°C, 50 sccm $N_2$ purge. 40°C, 10 min; 10°C/min to 150°C for 1 h or 3h; 10°C/min to 200°C for 30 min.

| bed # | sample | adsorbent | Time at 150°C, h | Wt% loss to 150°C | Wt% loss 150-200°C | total wt% | % of total to 150°C |
|---|---|---|---|---|---|---|---|
| 1 | 2012301-1 | ADS_A | 1 | 13.91 | 0.57 | 14.48 | 96.1 |
| 1 | 2012301-1 | ADS_A | 3 | 14.48 | 0.27 | 14.75 | 98.1 |

Based on the above data, all of the adsorbents tested are capable of adsorbing the organics, as indicated by the significant weight changes when exposed to the syngas. Several suitable candidates were found that were insensitive to sulfur accumulation (ADS_A and ADS_B), while others showed advantages of higher capacities and potentially easier regeneration (ADS_D and ADS_G).

## 7- Results from Campaign 3

The goal of Campaign 3 was to test thermal swing adsorption (TSA) approaches for regenerating the adsorbent. Testing was conducted during 20-25 May. The same gasifier setup was utilized, but the adsorption vessels were operated with a TSA cycle rather than PSA.

Adsorbents to be considered were determined from the results of the six-bed exposure testing described above and consisted of ADS_A, ADS_B, ADS_D and ADS_G. Four separate columns were packed with these materials as described in Table T3-14 and shipped to EERC's facility.

Table T3-14. Adsorbents packed in TSA beds.

| Acronym | Amount packed (g), feed end to product end |
|---|---|
| **ADS_A** | 6" glass beads, 3" ADS_C, 74" ADS_A |
| **ADS_B** | 6" glass beads, 3" ADS_C, 72" ADS_B |
| **ADS_D** | 9" glass beads, 71"ADS_D |
| **ADS_G** | 9" glass beads, 73" ADS_G |

The TSA columns were modified to permit more gas sample points near the feed end of the bed. Sample taps located 13" and 36" from the feed end of the adsorbent layer were typically used. Analysis of the breakthrough runs focused on benzene, since it was the most prevalent tar compound and was very easy for the tar GC to monitor.

The TSA cycle used in this work is described in Table T3-15. Each complete cycle took 12 hours to complete. External heaters on the vessels ramped the temperature at 2°C/min to a final temperature of either T1 (ADS_A, ADS_B, ADS_D) or T2 (ADS_G). Nitrogen was used to purge the vessels while heating, holding and cooling, and was also used to re-pressurize the columns before the feed step. Feed gas flow rate was 40 slpm, and the $N_2$ flow rate was 10 slpm.

Table T3-15. TSA Cycle Information.

| Steps | Time (min) | EOS Pressure (psig) |
|---|---|---|
| **Feed** | 360 | 400 |
| **Depress 1** | 1 | 130 |
| **Depress 2** | 1 | 20 |
| **Heat** | 75 | 5-10 |
| **Hold** | 101 | 5-10 |
| **Cool** | 180 | 5-10 |
| **Repress** | 2 | 400 |
| **total** | 720 (12 hrs) | |

The cycle schedule was tight, as we wanted to conduct at least five cycles with at least one of the columns, but still have time to look at the ADS_D and ADS_G adsorbents. An overview of the testing timeline is illustrated in Figure T3-22. The schedule was complicated by an unexpected gasifier shutdown in the middle of the test period.

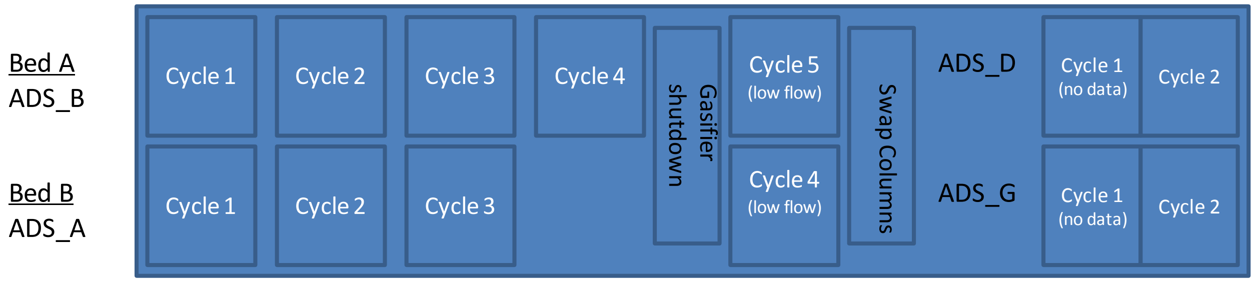


Figure T3-22. Overview of TSA Testing conducted in Campaign 3.

The goal of the first set of tests with ADS_A and ADS_B was to 1) evaluate the potential of these materials to remove the tar species, 2) determine if the adsorbent capacities were stable cycle to cycle, and 3) generate some useful design data.

Topic 2 was made somewhat more difficult by the fact that the level of tar components was not particularly stable during the test. During the first three TSA cycles the feed varies from about 100 to 200 ppm benzene. After the gasifier restart the benzene increases dramatically to above 400 ppm, then it returns back to 100-150 ppm for the last cycles. This must be considered when looking at the breakthrough results. The difference in capacity for 400 and 100 ppm benzene at 30°C based on some reference benzene/ADS_B isotherms is 2.4 versus 1.5. The ratio of retention times is then expected to be $t_{400\ ppm}/t_{100\ ppm}$ = (2.4/400) / (1.5/100) = 0.4. This means the breakthrough time for gas with 400 ppm benzene is expected to occur 60% earlier than a feed gas containing 100 ppm benzene. This could easily be mistaken as a dramatic reduction in the benzene adsorption capacity if the difference in feed level benzene is not recognized.

The feed gas tar composition throughout the campaign is illustrated in Figure T3-23. Organic species are present in roughly the following levels; 110 – 450 ppm benzene, ~1 ppm UN37, ~1 ppm toluene. $H_2S$ and $CO_2$ levels are at 0.27% and 57% and are stable throughout the test period.

An illustration of the raw organic species breakthrough at the 13" sample tap is presented in Figure T3-24 for cycle 4 of the ADS_B column. $H_2S$, $C_2H_6$, and $CO_2$ break through almost immediately. These components would have to be removed by the PSA rather than the TSA. Benzene is initially absent in the effluent gas. One ppm benzene breakthrough starts at around 60 min.

Breakthrough results for the first four cycles with the ADS_B column are plotted in Figure T3-25, and those for ADS_A are in Figure T3-26. The curves for different cycles are reasonably consistent.

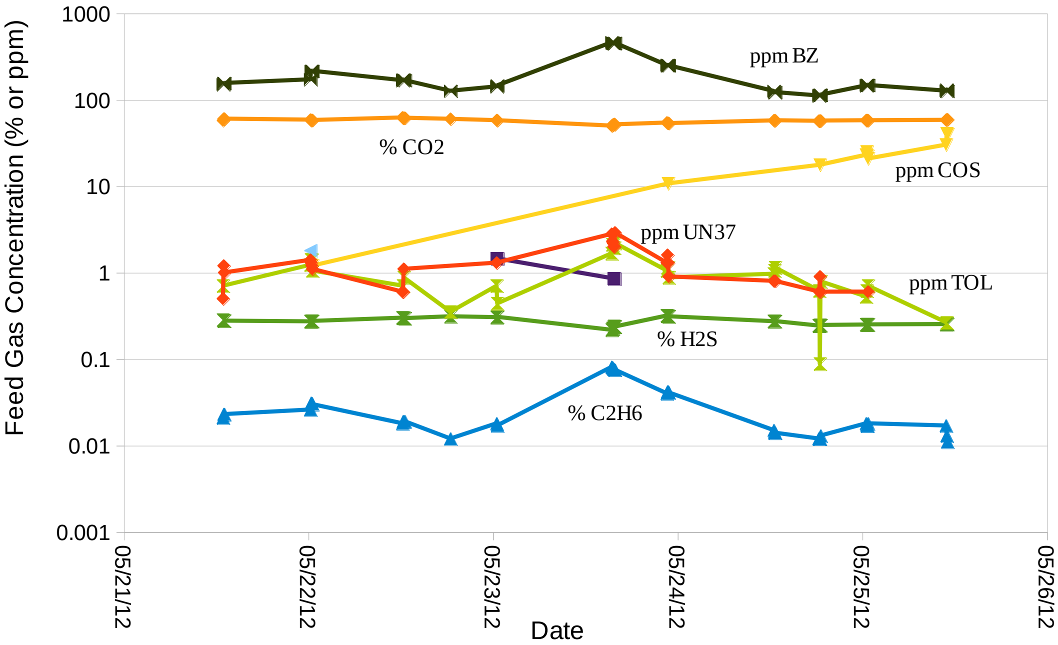


Figure T3-23. Composition of sour syngas feed to the sour PSA/TSA unit, Campaign 3.

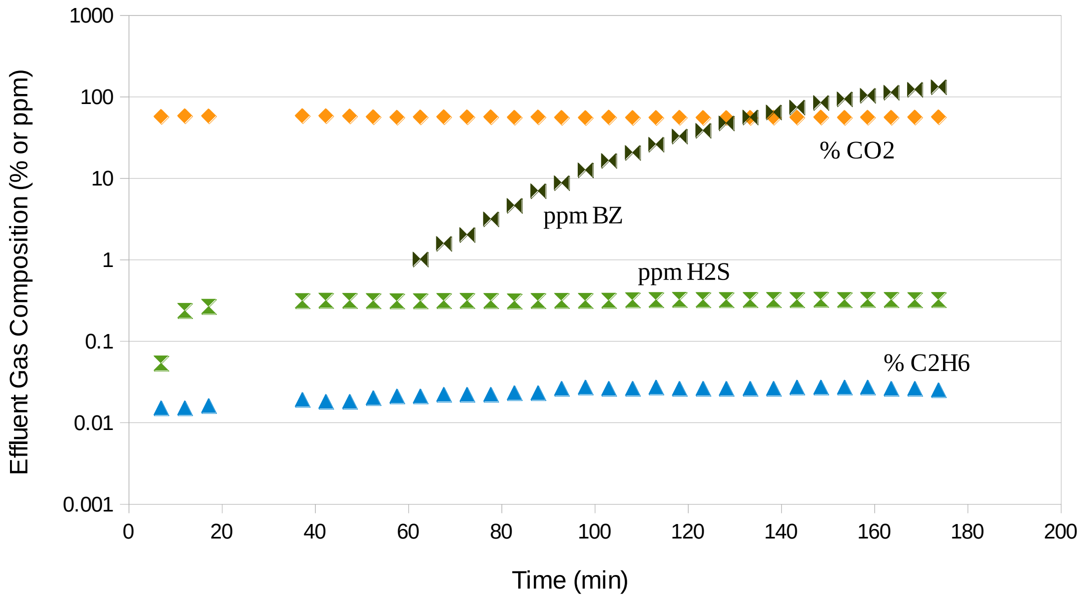


Figure T3-24. Effluent gas composition during 4th TSA test with ADS_B column.

Benzene adsorption capacities and length of the unused bed (or LUB) were determined from the 1ppm breakthrough time, the stoichiometric time, and an estimated feed gas benzene concentration. The calculated results are listed in Table T3-16. The benzene capacity is scattered around 3 wgt%. Increasing the benzene level in the feed increases the observed capacity, as expected. There is no indication that the adsorbent capacity is decreasing significantly from cycle to cycle. The adsorbent is capable of effectively removing benzene from the feed gas.

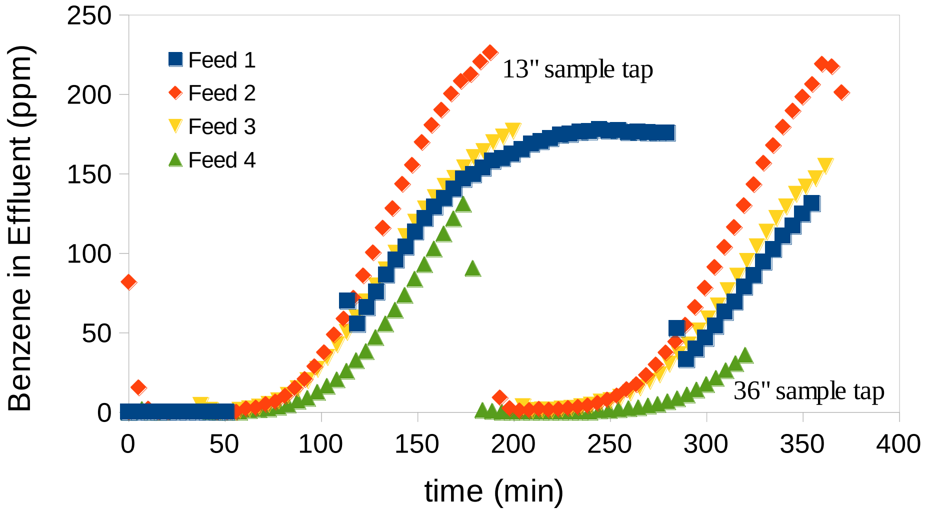


Figure T3-25. Multiple Benzene Breakthrough Curves for ADS_B.

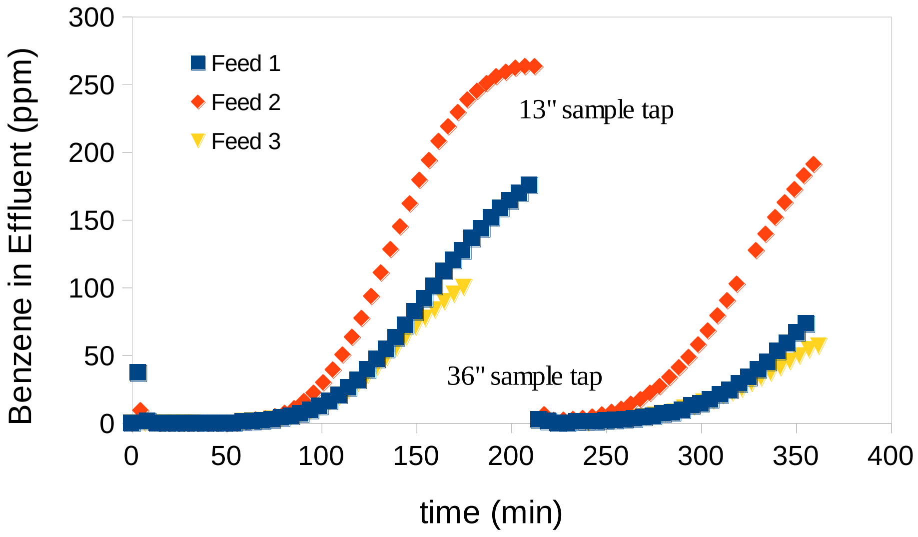


Figure T3-26. Multiple Benzene Breakthrough Curves for ADS_A.

Table T3-16. Summary of LUB's and Benzene Capacity from TSA Tests.

| Adsorbent | Cycle # | ppm Bz in feed | Sample tap position (in) | Relative N Bz | Relative LUB |
|---|---|---|---|---|---|
| **ADS_B** | 1 | 176 | 36 | 0.67 | NA |
| **ADS_B** | 2 | 220 | 36 | 0.82 | 0.97 |
| **ADS_B** | 3 | 170 | 36 | 0.64 | 0.97 |
| **ADS_B** | 4 | 140 | 36 | 0.57 | 0.77 |
| **ADS_B** | 5 | 344 | 13 | 1.36 | 0.50 |
| **ADS_A** | 1 | 200 | 36 | 0.84 | 0.98 |
| **ADS_A** | 2 | 265 | 36 | 1.00 | 1.00 |
| **ADS_A** | 3 | 136 | 36 | 0.56 | 0.99 |
| **ADS_A** | 4 | 230 | 13 | 1.07 | 0.56 |
| **ADS_D** | 2 | 148 | 13 | 4.19 | 0.32 |
| **ADS_G** | 2 | 130 | 13 | 3.22 | 0.17 |

Breakthrough curves for the ADS_D and ADS_G adsorbents are compared with the ADS_A and ADS_B data in Figure T3-27 (all at 40 slpm and 13 inch sample tap). Both the ADS_D and ADS_G adsorbents have much higher benzene capacities and therefore longer breakthrough times. The ADS_G adsorbent also appears to have rather fast mass transfer indicated by the steep mass transfer zone. These observations are supported by the relative capacities and LUBs listed in Table T3-16. These two materials are potentially more attractive adsorbents for removing tar species, although more testing is needed to make sure there is no capacity decline with cycling.

At the end of Campaign 3, the TSA beds were regenerated, sealed, removed for the test rig and sent to APCI for post mortem analysis. Samples were taken from the inlet and outlet ends of the four used beds and evaluated via TGA. Weight losses to 200°C were quite small, consistent with the fact that the beds had been regenerated before shipping. Only $H_2O$ and trace or no $CO_2$ were detected, and most of the sample weight losses can be attributed to water. No hydrocarbons were observed. Changes in $CO_2$ capacities relative to the fresh adsorbent were quite modest (< 15%). Based on these results, use in a TSA process resulted in no significant change in adsorbent properties and no organic species were retained on the adsorbents.

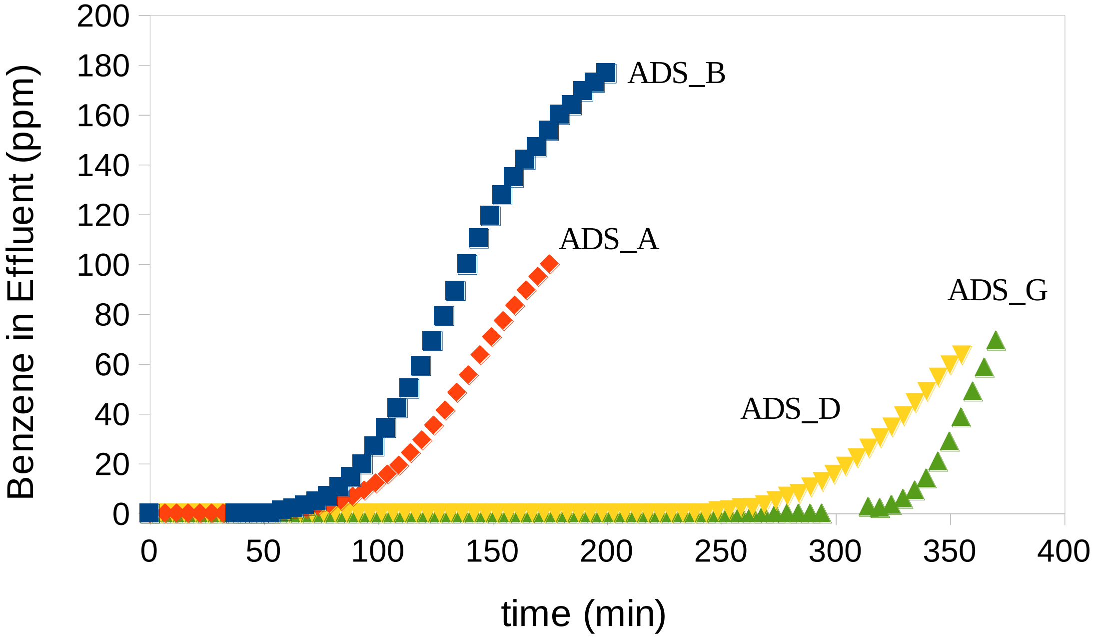


Figure T3-27. Benzene Breakthrough Curves for Adsorbents ADS_B, ADS_A, ADS_D, and ADS_G.

# Task 4. Design pilot unit:

The goals of this task were to determine what type of sour PSA or sour TSA/PSA process was most attractive, evaluate the impact of sour feed gas conditions on PSA materials of construction, and use the generated experimental process data to validate a model for predicting industrial and pilot-scale performance.

## 1- Identify best adsorption process and generate heat and mass balance flow sheet:

There are two ways to implement sour PSA technology with low rank PRB coal feedstock. The first is to utilize a fluidized bed gasifier to generate a tar-laden sour syngas, pass that gas to a TSA unit for removal of the organic tar components, and then introduce the tar-free sour syngas to the sour PSA unit for rejection of $H_2S$ and $CO_2$. This requires additional TSA vessels, adsorbent, valving, and thermal regeneration equipment. The second approach is to eliminate, or greatly reduce, tar formation by conducting the coal gasification in a high temperature gasifier, e.g, an entrained flow gasifier. These are known to be effective for handling PRB coal, and yield essentially tar-free syngas due to the more aggressive reaction conditions which facilitate the decomposition of aromatic species. Entrained flow gasifier designs make up a sizable portion of the large scale gasification facilities being constructed today.

The clear advantage of the latter approach is obvious, as the installation and operating cost of the TSA system is eliminated. This reduces the overall technical risk in the pilot while still being applicable to the commercial scale market. The pilot could also be applied to fluidized bed gasification processes where the use of a TSA would have to be evaluated. As with all gasification projects, a number of factors go into the overall cost of gas produced from the gasification plant and would therefore influence the cost of a Sour PSA pilot test. For the purposes of this study, we assumed that a high temperature gasifier was chosen and the tar content of the feed gas entering the sour PSA was negligible. The gases to be purified are considered a slipstream in an existing process, with product gas produced from the Sour PSA pilot either being flared or used as useful product and the tailgas from the unit being processed further in the existing acid gas removal process. This minimizes the requirements of the pilot while still yielding the needed information for commercial development.

## 2- Materials of Construction

Sour syngas can be corrosive, especially when conditions are right for water in the gas to condense on the inside surfaces of piping and equipment. Corrosivity is brought about via the formation of carbonic acid and hydrosulphuric acid. The National Association of Corrosion Engineers (NACE) has defined a $H_2S$ partial pressure of 0.05 psia which triggers sulfur-based corrosion consideration.

Feed gas to the sour PSA system contains sulfur components and $CO_2$ at levels of around 0.3 – 3% and 20-50%, respectively, depending on the solid feed stock and application. $CO_2$ in the feed gas can lead to corrosion of carbon steel when it is in contact with standing water. Thus, any areas where condensation and collection of water are possible would require stainless steel construction. There are no areas in the sour PSA system where condensation conditions are expected. However, condenser/knockout units preceding the sour PSA boundary must be evaluated.

The $H_2S$ partial pressure in the feed is around 1-13 psia (assuming 415 psia total pressure), so it is clearly above the NACE criteria. $H_2S$ exposure generally leads to metal cracking rather than corrosion, and has greatest impact on hardened steels. Carbon steel pipe is immune to $H_2S$. Carbon steel plate material is not, though. Since PSA vessels and the purge tanks would be fabricated from plate steel, they need to be post weld heat treated (PWHT) to reduce the potential for $H_2S$ stress cracking. The impact of PWHT on overall cost is relatively small.

Materials of construction must also be considered for valves. Switching valves constructed from stainless steel would be acceptable for $H_2S$ service. Components which are additionally heat treated for hardening need to be investigated further, as this process may make them more susceptible to $H_2S$ corrosion.

Valves on the product side of the PSA system will see $H_2S$ levels around 2-10 ppm. The partial pressure of $H_2S$ is around 0.004 psia, which is lower than the 0.05 psia NACE criteria. The $H_2S$ level would need to reach 120 ppm at 415 psia to trigger NACE concerns. Levels this high are not anticipated.

Valve skids/piping would not need PWHT due to their fabrication from carbon steel pipe rather than plate steel.

Soft sealing materials that are compatible with sour syngas include Kel-F, PTFE, FEP, Teflon, Tefzel, Kynar, PVC, Polycarbonate, Kalrez, Buna-N, Neoprene, and Polyurethane.

### 3- Process design and cost estimate for pilot unit:

In this effort, our in-house adsorption process simulator was validated by comparing simulation results with experimental sour PSA cycle data. First, though, the equilibrium and mass transfer properties of the various components of the sour syngas feed gas had to be characterized. This involved matching experimental isotherm data with an isotherm model, and then fitting the simulation results to breakthrough data to refine the mass transfer parameters. Both of these tasks had been previously completed before the DOE project started. The isotherm and mass transfer parameters were simply reviewed and updated in this work.

An example of the fit of experimental isotherm data for $H_2S$ and $CO_2$ with the isotherm model is illustrated in Figure T4-1 and T4-2. The model does a good job of describing the data at both 30 and 70$^{o}$C.

A comparison of experimental $H_2S$ breakthrough data with predictions from the adsorption process model is illustrated in Figure T4-3. The data were obtained with a feed gas containing 2.3% $H_2S$ in $H_2$ at 400 psig and 30$^{o}$C. Figure T4-4 shows the bed temperature measured at various positions along the column during the breakthrough test. Exotherms approaching 20$^{o}$C are observed. Since the equilibrium properties are rather strong functions of temperature, it is important to effectively model the heat transfer properly so the simulations give similar temperature profiles. This is the case in the current case, as indicated by the relatively good agreement with the results from the adsorption process simulator (lines in the figure).

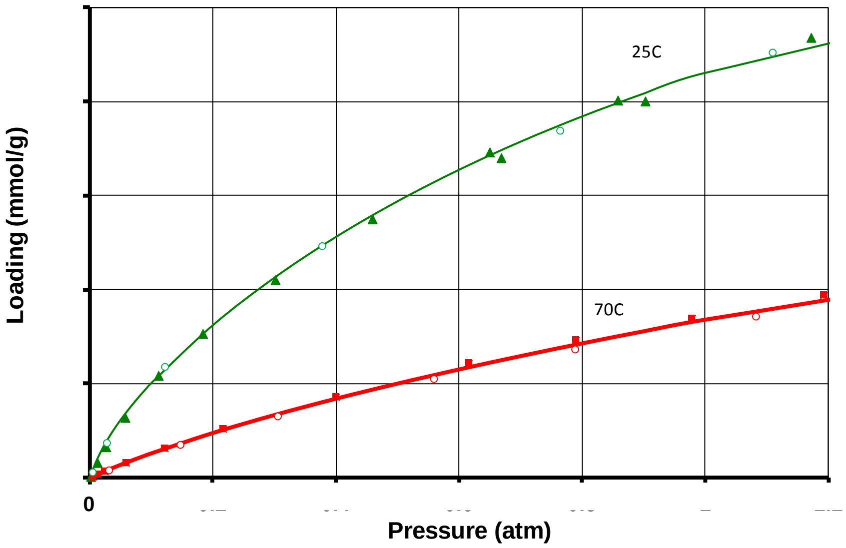


Figure T4-1. Isotherm for $H_2S$ on adsorbent ADS_A; symbols - experimental data, lines - isotherm model.

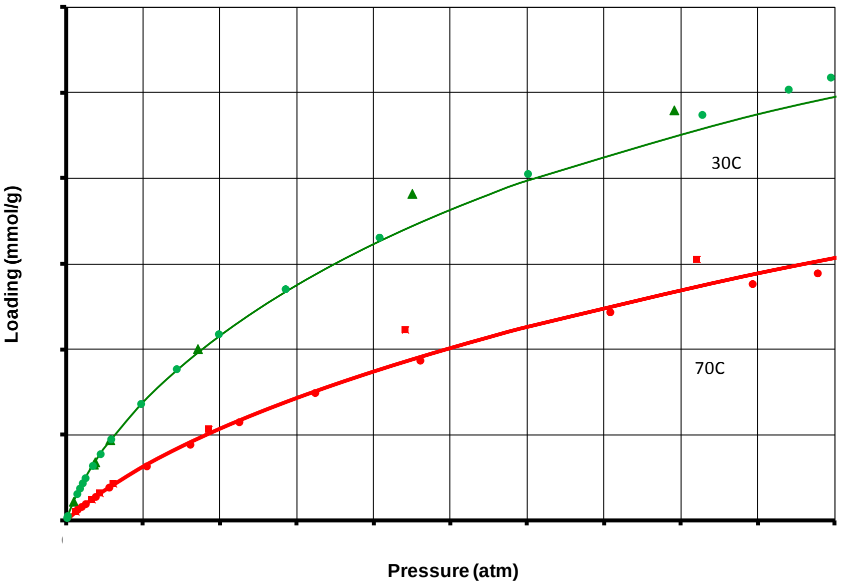


Figure T4-2. Isotherm for $CO_2$ on adsorbent ADS_A; symbols - experimental data, lines - isotherm model.

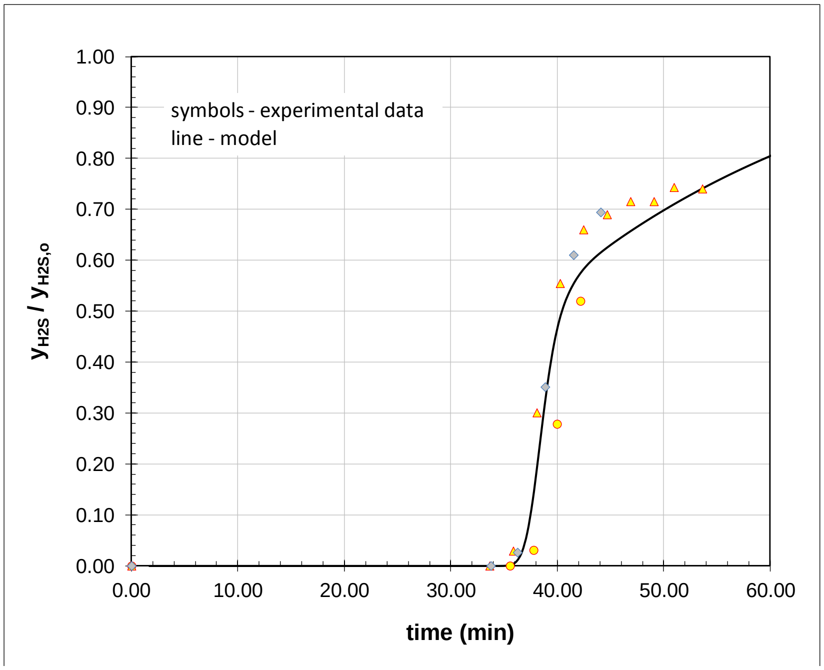


Figure T4-3. Comparison of $H_2S$ breakthrough data with simulation results.

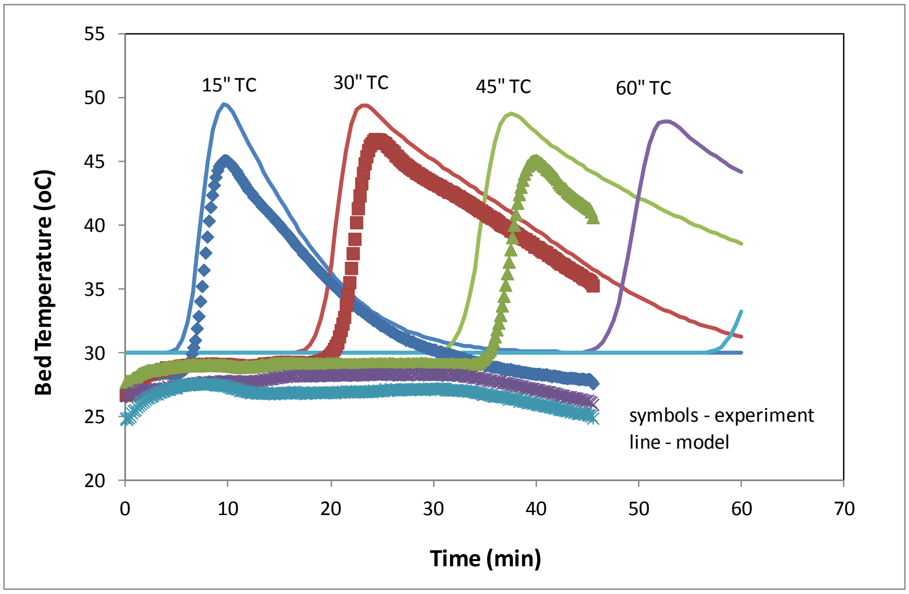
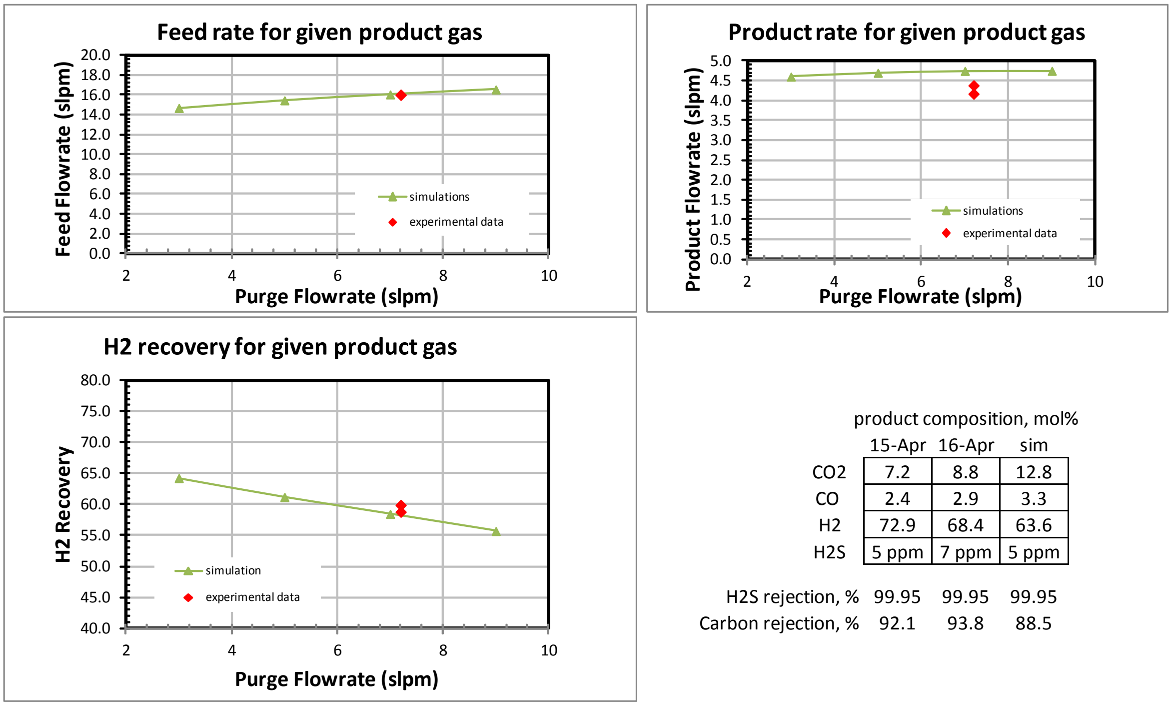


Figure T4-4. Comparison of bed temperature data and simulation predictions.

The location of the breakthrough curve in time is largely due to the isotherm model parameters, and the shape of the breakthrough curve is controlled by the mass transfer model/parameters and heat transfer properties of the adsorption column. Figure T4-3 shows that both the midpoint and shape of the breakthrough curve is well predicted by the model.

Once the equilibrium, mass and heat transfer parameter set had been developed, the simulation model was used to predict the performance of the Sour PSA unit when operated under cyclic conditions. The data for Campaign 2 were chosen for this comparison (see Figure T3-13), and results are plotted as red symbols in Figure T4-5. Observed $H_2S$ levels were 5 and 7 ppm. Simulations were conducted by varying the purge rate over a specific range while adjusting the feed flow rate to yield product gas containing 5 ppm $H_2S$. The simulations results are plotted as lines in the figure. The $H_2$ recovery is slightly under-predicted by the process simulator, while the overall carbon capture is significantly under-predicted (88% predicted, 93% observed). There is more slip of $CO_2$ to the product gas in the simulations, which yields the lower carbon recovery prediction. Thus, the simulations are a conservative estimate of cyclic performance, and this bias will likely hold when the model is used to predict industrial scale performance. Nevertheless, the simulator is shown to be an effective tool for characterizing the lab-scale PSA data, and was used to predict industrial-scale PSA performance for a much more complicated process cycle.



product composition, mol%

| | 15-Apr | 16-Apr | sim |
|---|---|---|---|
| CO2 | 7.2 | 8.8 | 12.8 |
| CO | 2.4 | 2.9 | 3.3 |
| H2 | 72.9 | 68.4 | 63.6 |
| H2S | 5 ppm | 7 ppm | 5 ppm |
| H2S rejection, % | 99.95 | 99.95 | 99.95 |
| Carbon rejection, % | 92.1 | 93.8 | 88.5 |

Figure T4-5. Comparison of sour PSA process performance with predictions from the process simulator.

### 4- Industrial sour PSA design

The simulator was used to design an industrial-scale sour PSA process. The sour feed gas was defined by the output of the ASPEN simulations described in Task 5 and was processed at 35 atm. The PSA unit was targeted to achieve 3 ppm $H_2S$ or less in the product gas. Only a single layer of adsorbent was considered in the PSA unit. This layer effectively rejects $H_2S$ and most of the $CO_2$, but $CH_4$ and CO mostly pass through with the $H_2$. This means that the degree of carbon capture across the PSA must consider CO and $CH_4$ as well as $CO_2$. The unit was designed to achieve at least 92% overall carbon capture. The $H_2$ recovery was greater than 90%. This much higher level of recovery (than experiments in Task 3) was attained through use of multiple pressure equalization steps. Each bed executed a specific set of process steps, and the total cycle time was 10 minutes.

A simple schematic of the sour PSA system is presented in Figure T4-6. One train of the sour PSA system consisted of ten packed beds with dimensions equivalent to available commercial units. A valve skid is piped to the vessels, and the process steps are executed by opening/closing the valves via a plant process control system. Valves include on/off switching valves as well as some throttling valves for flow control. Effluent gas during blowdown and purge steps is collected in a large vertical purge tank which dampens flow and concentration variations. Three PSA trains are needed to handle the total sour syngas flow in the case study of Task 5.

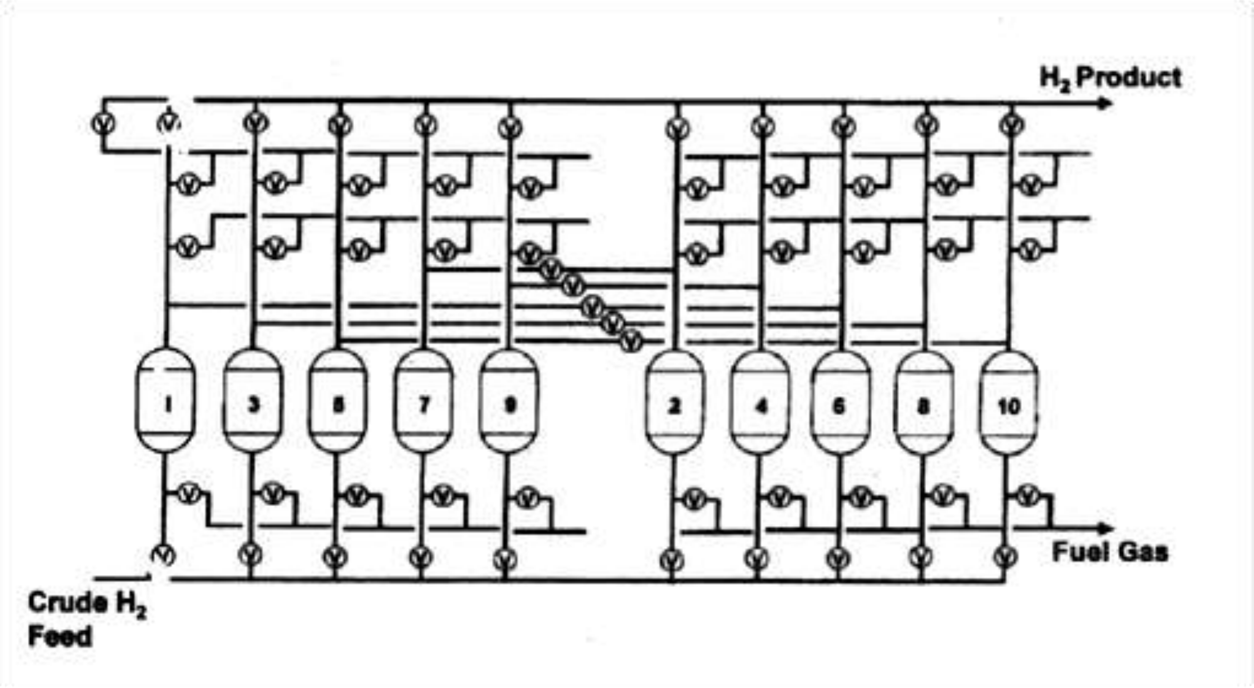


Figure T4-6. Simple schematic of one train of the sour PSA process.

The PSA cycle and process equipment is similar to that used by Air Products to produce $H_2$ sweet syngas streams. Thus, there is a great deal of familiarity with design, operation, and cost of these items. The production of these systems has been optimized over the years, and currently all of the components are shop fabricated. This reduces fabrication costs and simplifies installation efforts. Our in-house capital cost evaluation software was used to generate an overall capital cost that included costs for the PSA vessels and tanks, adsorbent, automated and manual valves (skidded), process piping, project management and engineering labor, and freight.

## 5- Sour PSA Pilot scale process design

Further development of this technology includes demonstration of performance in a multi-bed (i.e., 10 bed) process unit capable of running the more complex industrial process cycle. It must process real sour syngas, and should be capable of running continuously for at least three months of time. This suggests that a pilot unit should be designed and constructed near an industrial gasifier. A sour slip stream could then be easily fed to the sour PSA. The product gas stream from the PSA would be obtained at pressure, so it could potentially be reintroduced to the downstream gasification system. The low pressure waste gas from the sour PSA would contain substantial $H_2S$ and $CO_2$, so it could not be simply vented. Ideally it would be worked into the existing gasification train's acid gas removal unit.

The pilot sour PSA unit could then consist of a smaller version of the industrial PSA described above. Adsorber vessel length would be kept the same as the industrial design. The adsorber vessel diameter is an important parameter: it defines how much syngas feed is required and has a strong influence on total cost. Generally, a smaller diameter is chosen, but this can lead to difficulty in interpreting experimental results due to the impact of heat loss. There is therefore a tradeoff with respect to column diameter that was taken into account into the vessel design for the pilot.

The pilot sour PSA system would continuously process a flow of roughly 200 lbmole/hr (1.9 MM SCFD) of feed.

The same capital cost estimation software was used to determine the cost of the smaller system. An additional $150 K was included to account for the additional analysis equipment (GC, sulfur analyzer, thermocouples in each vessel, additional flow meters). A compressor to re-inject the waste gas to the gasifier sulfur loop was not included as it requires more detailed site-specific input. An installation factor was applied that was consistent with Air Products' experience. On this basis, the estimated cost for the pilot-scale sour PSA unit is $MM12.

The nature of the gasification site will determine if the sour syngas requires cleanup before being sent to the PSA. A tar-containing gas would require a front-end TSA unit to reject the tars. This would consist of a two-bed system capable of being thermally regenerated to at least 200°C. Regeneration gas would consist of plant $N_2$ heated with an electrical heater.

A plot diagram of the proposed sour PSA system is illustrated in Figure T4-5.

Figure T4-5. Proposed Sour PSA plot plan.

# Task 5: Techno-Economic evaluation of sour PSA technology:

We have completed a preliminary techno-economic assessment of the impact on Cost of Electricity (COE) from incorporation of Air Products' Sour PSA technology into a base IGCC power plant with $CO_2$ capture design utilizing low rank coal.

The recent report by NETL ("Cost and Performance Baseline for Fossil Energy Plant: Low Rank Coal to Electricity: IGCC Cases") was used to guide the methodology of estimating the overall economics of the project. The methodology was also applied to the information provided for the base line reference plant case to insure the accuracy of our approach. Upon review of the available cases, we selected the Siemens gasifier with $CO_2$ capture and sub-bituminous PRB coal case (referred as S3B case in the NETL report) as the baseline case for our assessment. The reference system, as shown in Figure T5-1, includes a coal handling system that feeds a Siemens gasifier, an air separation unit (ASU) providing $O_2$ to the gasifier as well as $N_2$ for diluting the gas turbine fuel. The raw syngas produced by the gasifier is scrubbed and then shifted to $H_2$ and $CO_2$ prior to Hg removal and being 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.

For our evaluation, we chose to keep the gasifier size fixed but replaced the AGR unit with Air Products Sour PSA technology. This results in the resizing of 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 modification. This reduces the errors introduced with having to scale multiple unit operations costs. However, this does introduce issues with the gas turbines 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 as well as 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 (for a given volumetric heat value of the fuel) performance as the reference case. This is a reasonable assumption as the gas turbine is operated near or below the reference case fuel conditions.

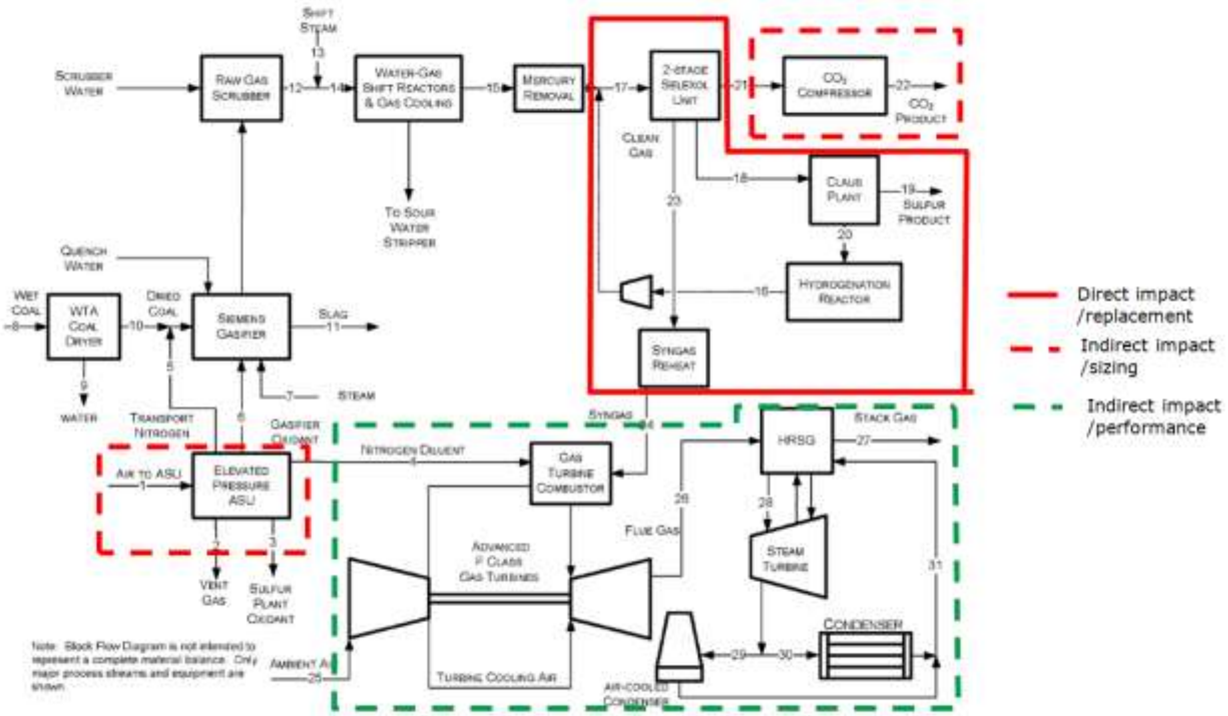


Figure T5-1: process and equipment impact

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, Initial and Annual O&M Costs, allowing for a simple estimation of the COE. Process simulations are used to estimate equipment size and equipments costs are estimated based on exponential factors and process parameters used in the NETL report and provided to Air Products by NETL. Note that the economic model is capable of calculated COE from a cash flow analysis similar to what the DOE's Power System Financial Model (PSFM) does. Estimating COE from the cash flow analysis gives consistent results with those published by NETL for non COE capture case. However, the COE published by NETL for the $CO_2$ capture cases do 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. In order to insure consistency, in this study we calculate the COE 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, we also calculated a COE number for the Sour PSA system based on the cash flow statement. The COE calculations were completed for two Sour PSA process configurations using different options for Sulfur disposition.

## 1- Process Selection:

The results from the experimental tests conducted in Task 3 provided a basis for which to model, design, and cost two process design configurations for Task 5 as well as a pilot unit for Tasks 4. 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 capture of $CO_2$ and other criteria pollutants (e.g., SOx, NOx) 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 T5-2. The system consists of three components: i) a PSA unit for purification of the sour syngas stream; ii) sulfur treatment, and iii) $CO_2$ polishing and compression. Unlike 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).

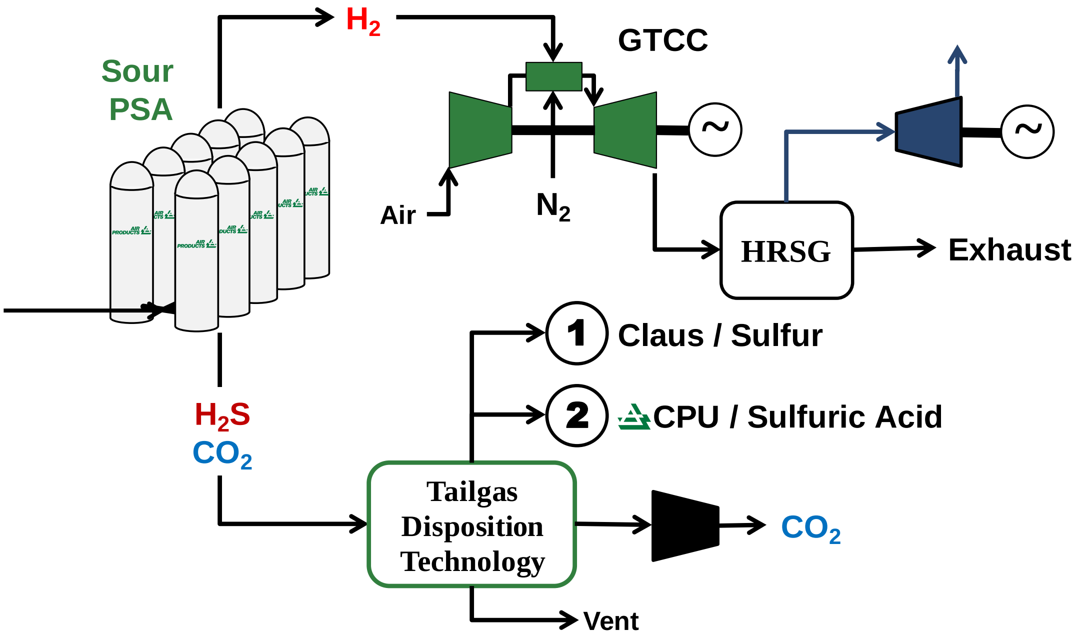


Figure T5-2: Tail gas Treatment Options for Power Applications of Sour PSA Technology

One option for a Sour PSA process configuration is the Sour PSA unit followed by a sour oxy-combustion unit and finally the $CO_2$ purification / compression unit (CPU), Figure T5-3. The Sour PSA is fed sour syngas and produces high-pressure $H_2$-enriched product and 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.

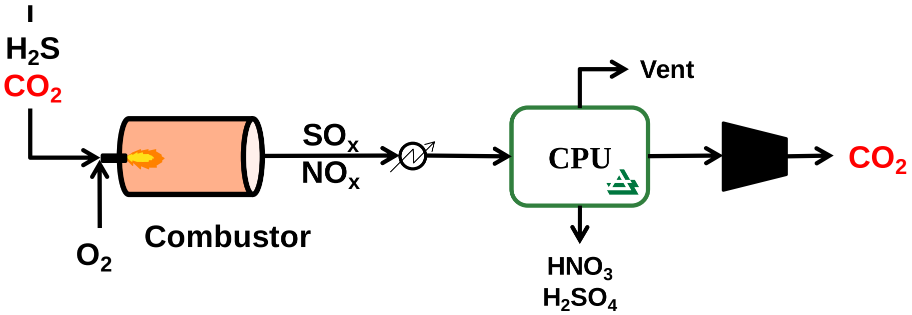


Figure T5-3: The Combustion/CPU tail gas treatment option

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 SOx-free, NOx-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 in an appropriate manner. The specialized $CO_2$ purification / compression system, which includes $SO_2$, NOx, and inert removal systems, was originally developed by Air Products for oxyfuel $CO_2$ capture for pulverized coal combustion power boilers and is currently under the pilot stage of development at Vattenfall's Oxyfuel Combustion Pilot 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 T5-4 shows a schematic in which $H_2S$ and a small portion of $CO_2$ is 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 have specialized sterically hindered amines such as ExxonMobil's Flexsorb®. The acid gas produced from the stripper of the AGE is sent to the Claus plant. Process parameters and performance were estimated based on existing information datasets. The sweetened gas then is compressed to ~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 lights are at pressure and rich in hydrogen and other combustibles originally rejected in the tail gas stream with low levels of $H_2S$ and $CO_2$. The stream is 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.

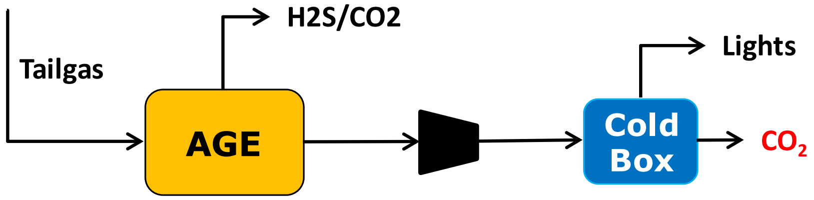


Figure T5-4: Sour PSA Acid gas enrichment tail gas treatment option for elemental sulfur product disposition followed by a cold box to purify the product $CO_2$

## 2- Process Simulation:

The experimental results obtained at EERC were used to simulate the performance of the PSA system using Air Products' proprietary simulation software, SIMPAC. The simulation results, in combination with our 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. The sour PSA design consists of 3 trains of 10 beds with dimensions equivalent to available commercial 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 $CO_2$, 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.

We investigated (process simulation and economic assessment) two different processes for disposition of the Sulfur in the PSA tail gas. The oxy-combustion of the tail gas followed by $CO_2$ purification / compression to convert Sulfur into sulfuric acid, and an Acid Gas Enrichment technology supplying an $H_2S$ rich stream to a Claus Plant process to dispose of the Sulfur in its elemental form. These options were briefly outlined above and will now be examined in context of the two comparing cases.

### *Reference Cycle*:

In order to compare the performance of Air Products' Sour PSA technology to the reference case using Selexol as described in the NETL report (S3B case), 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 that we can compare the different cases on the same basis we have simulated the process of the power island for the reference case with some assumptions. Our assumptions are summarized in figure T5-6 that shows a simplified PFD of the process we have simulated. The corresponding heat and mass balance summary is available in table T5-a1 in the appendix.

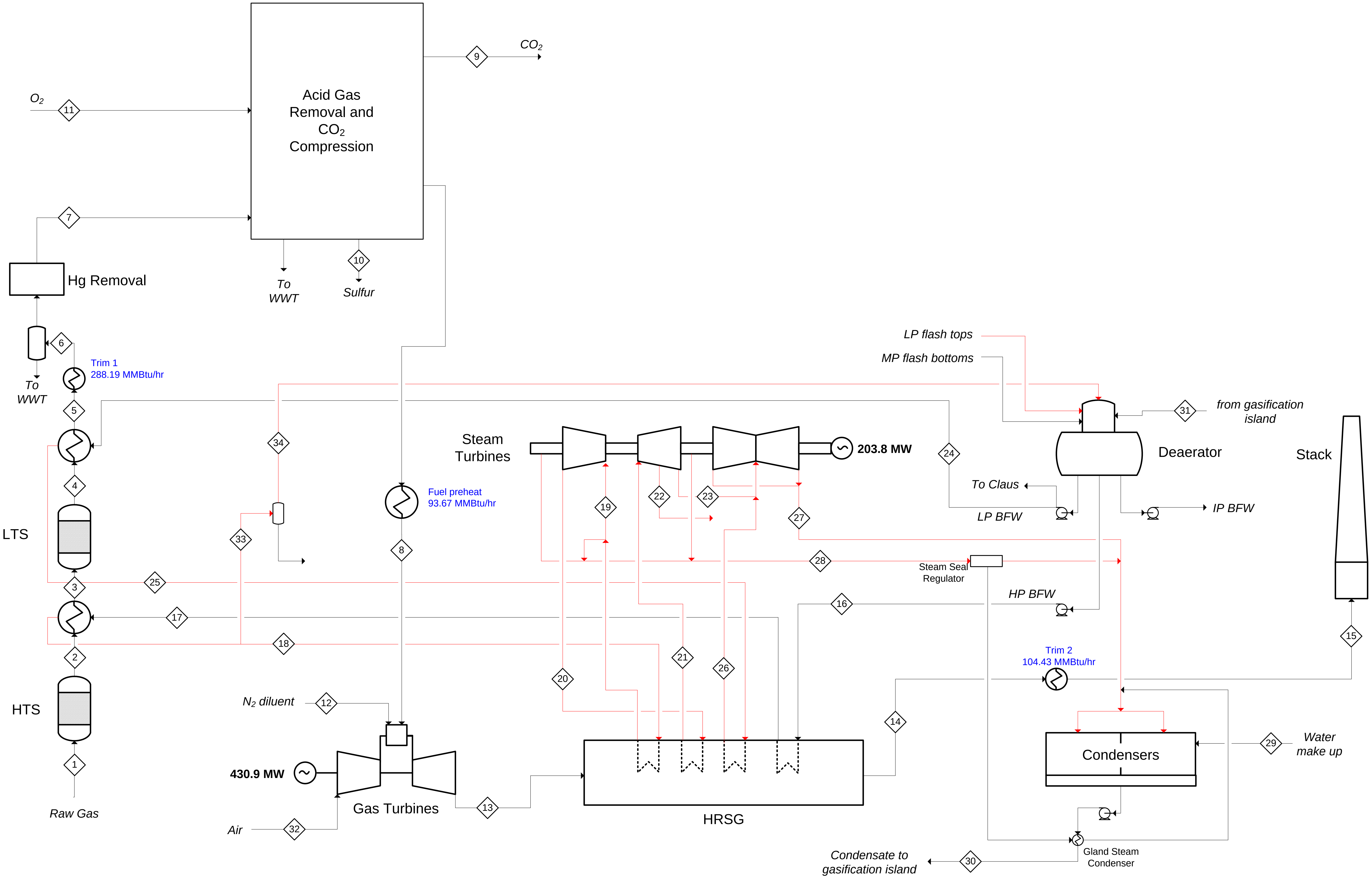


Figure T5-6: Simplified PFD of the simulated power island of the S3B reference case based on DOE/NETL report 2010/1399 p339

Comparison of our simulated process with the available information from the NETL report suggests that there is, in our simulated process, some heat left over in the raw syngas stream 5 and in the HRSG exhaust stream 14. These are represented on figure T5-6 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, our simulation suggests that there is about 288 MMBtu/hr of left over heat between 315 F and 95 F available on the raw syngas. We estimate that this heat is more than what is necessary to supply the regenerating heat duty for the Selexol unit. In addition, there is about 104 MMBtu/hr of left over 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 too. Fuel preheat to 420 F (per NETL report) requires about 94 MMBtu/hr that are not explicitly reported in NETL study. We therefore make the assumption that there is enough heat left over in the HRSG to provide the heat duty necessary for the fuel preheat.

### ***Oxy-combustion of sour PSA tail gas***:

This sulfur disposition process relies on Air Product's technology and involves combustion in pure oxygen (oxy-combustion) of the tail gas, followed sour gas compression, see Figure T5-3. The oxy-combustion results into SOx formation that is converted into sulfuric acid during the sour compression step. Such a process allows the Sulfur contained in the sour PSA tail gas to be converted into Sulfuric acid resulting in a clean $CO_2$ product stream. The technologies are both in development and carry with them higher degrees of uncertainty. Work unrelated to this project has been conducted to reduce this technical risk. Specifically the work on the compression and purification unit (CPU) completed under DOE award DE-NT0005309 entitled "Flue Gas Purification Utilizing SOx/NOx Reactions during Compression of $CO_2$ Derived from Oxyfuel Combustion" as well as work completed at our pilot site at Vatenfall's Oxyfuel Combustion Pilot at Schwarze Pumpe Germany. Initial testing was conducted in our combustion labs to provide the basis of our oxy-combustion process step. The initial experimental information, plus our company's expertise in oxyfuel combustion, is sufficient for this level of analysis. The next step in development would be required at a process pilot or demonstration unit.

The implementation of the Oxy-combustion of the sour PSA tail gas in the coal gasification to power process that we have simulated is described on the simplified PFD provided in Figure T5-7.

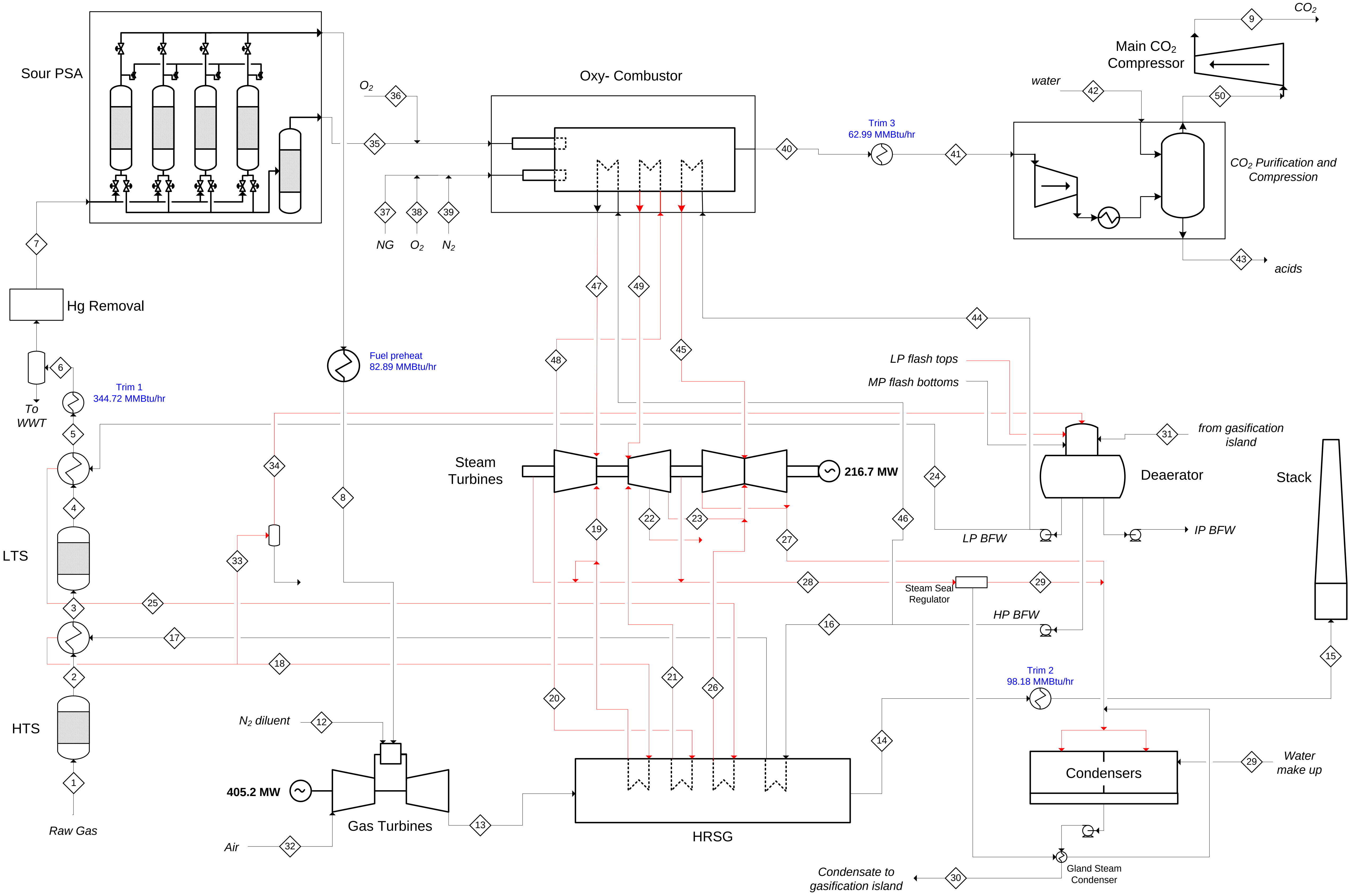


Figure T5-7: Simplified PFD of the Sour PSA with Oxy-combustion of tail gas Plant (adapted from DOE/NETL report 2010/1399 p339)

The PSA tail gas is primarily $CO_2$ and has a low heating value. A natural gas fired pilot burner is therefore used to insure the temperature in the furnace exceeds the auto-ignition temperature of the tail gas and therefore complete combustion of the fuel. In order to limit the volume of flue gas to be treated and reduce the $N_2$ introduced into the flue gas, an oxyfuel pilot burner was employed. Recommendations given in NFPA85 for igniters in continuous service were used and specified the design of the pilot burner to be 10 % of the full load burner. This resulted in using about 33.5 MMBtu/hr (HHV) of natural gas as auxiliary fuel in the process. Note that this is probably a conservative approach as NFPA85 uses air/fuel pilot burners as a reference point where our approach uses a much hotter, more stable oxyfuel pilot. Air Products' sour compression technology requires that the feed gas (in this case, the flue gas of the sour gas oxy-burner) falls with a certain SOx/NOx ratio range to produce an economic design. As mentioned previously, NOx acts as a catalyst to convert SOx to sulfuric acid. Based on Air Product experience, a small amount of low pressure waste $N_2$ (37 lbmol/hr) was added to the system to meet the downstream NOx requirement.

The exhaust stream of the sour gas oxy-combustion system is slightly above atmospheric pressure and at elevated temperature. In order to be fed into the sour compression unit for SOx removal, the stream needs to be cooled down to about 155 F (slightly above dew point). This represents about 323 MMBtu/hr available to raise steam. Not all the energy in the system was used to generate steam however. We took the conservative approach of producing steam at the same HP, IP and LP conditions and mass flow ratio that the power producing steam turbine in the NETL base case uses to avoid any configuration change in the power island. The exhaust gas of the sour gas oxy-combustion system is used to raise steam in the same way the HRSG is used to raise the steam for the steam turbine in the reference case. High pressure boiler feed water is pumped to make 108188 lb/hr of HP steam (2266 PSIA, 992 F) that is sent to the HP stages of the steam turbine, then extracted at 492 PSIA, 618 F, reheated to 1000 F, sent to the IP stages of the steam turbine. An additional 107977 lb/hr of LP steam (65 PSIA, 409 F) is generated and sent to the LP stages of the steam turbine. Note that we set up the simulation so that the steam leaks and steam export are proportional to the total HP, IP and LP steam raised in the HRSG and the oxy-combustor boiler. In these conditions, the steam raised by cooling the exhaust of the sour gas oxy-combustion system allows to generate 30.3 MW of electric power. However, because of constraints on the steam level and conditions we assume (in order to fit the reference steam cycle design) not all the heat contained in the sour gas oxy-combustion system exhaust stream can be used since it would result in temperature crossover. Indeed, with a 40 F minimum temperature approach constrain in the steam generator heat exchanger, our simulation shows that the exhaust gas is cooled down to only 321 F. This means that only about 80% (260 MMBtu/hr) of the available heat is used and therefore about 63 MMBtu/hr of heat, available between 321 F and the required 155 F feed temperature for the sour compression unit, is rejected (see Trim 3 on figure T5-7). We can't find use for that heat within the design constraints imposed by the existing steam cycle as provided by the reference model, but we believe that it could be used to increase overall power production and efficiency if that constraint was lifted. In other words, the amount of power we estimate to be able to raise from the heat of the sour gas oxy-combustion system exhaust is conservative.

The cooled exhaust of the oxy-combustion system is then sent to the $CO_2$ purification and compression unit (CPU) and compressed up to 435 PSIA in a multistage compressor. This unit uses about 31 MW of parasitic electrical power. A water contacting reaction column, converts the SOx and NOx to sulfuric and nitric acids respectively. For this particular study, the system was designed to convert ~98% of SOx and ~90% of NOx, effectively purifying the exhaust $CO_2$ stream to having less than 122 ppm of SOx. The SOx and NOx parameter for the CPU design were chosen to match the needs for this study but do not set a limit for what the technology can achieve. The resulting $CO_2$ product stream is at 435 PSIA and composed of 95.9% $CO_2$, 2% $N_2$, 1.5% $O_2$ with other low level impurities and is ready for drying and further compression to pipeline pressure (2215 PSIA). Because of the $N_2$, $O_2$ and Ar level, the $CO_2$ purity does not meet the suggested NETL specification but several commercially available technologies could be used to achieve this. At this stage we have not considered any technology to further clean the $CO_2$ stream as we did not understand the basis in which the purity requirement was set nor was the specification an absolute requirement. The overall carbon capture rate is 93.4% when coal and natural gas inputs are included (95.6% $CO_2$ capture).

By maintaining the gasifier output at the reference case conditions, the use of the sour PSA technology results in a 5.8% reduction in the amount of $H_2$ being sent to the gas turbines. The fuel sent to the gas turbines is obtained by diluting the sour PSA product with $N_2$ in order to achieve a fuel with a LHV of 120 Btu/SCF, consistent with the reference case. As a consequence of the lower amount of $H_2$ available, the gas turbines total output is reduced by 5.9% to 405.2 MW (assuming same heat rate as the reference plant). Accordingly, we model the effect of the size reduction on the gas turbines on the HRSG duty and steam generation. Our approach is to pass the effect of the reduced heat available from the exhaust of the gas turbines entirely on the power steam generation, leaving all the process steam cycle unchanged with respect to the reference cycle. Again, this results in an 8.5% reduction in the amount of steam that can be generated in the HRSG. This reduction is partially compensated by the additional steam generated by the sour PSA tail gas oxy-combustor boiler. However, the constraints on the HRSG/Steam turbine arrangement leave significant heat on the table which could be used to produce additional MW. Similarly to the reference case (see above), the 98 MMBtu/hr (Trim 2 on figure T5-7) of heat left over in the HRSG flue gas is enough to preheat the gas turbines fuel (only about 83 MMBtu/hr). However, as shown on figure T5-7, there is still 345 MMBtu/hr available in the raw syngas (Trim1) and 63 MMBtu/hr available in the oxy-combustor boiler exhaust (Trim 3). In the reference case, some of the heat from the raw syngas stream was used for the heat duty of the $H_2S$ stripping column of the Selexol unit. The Sour PSA with oxy-combustion of the tail gas would not have the same constraint, and thus a steam cycle utilizing this lower level heat would result in additional MW of power production. The increase in power production from the steam cycle was not analyzed as this requires a re-optimization of the steam cycle. Ultimately, the steam turbine may change as a result of that analysis. We did not consider this under the current time and budget constraints of the project. The summary of the heat and mass balance for the simulated process is available in table T5-a2 in the Appendix.

As reported in a previous published work, the oxy-combustion option lead to an addition of ~4% more net power from the plant compared to the baseline case. This is partly due to the higher levels of sulfur in the fuel (petroleum coke was the feedstock) which result in higher $H_2S$ and more heat released in the combustion step, along with a more conservative approach to the amount of additional natural gas being used as a pilot in that case. The additional constraints of the HRSG arrangement to match the baseline case makes up for the major differences between previous results and the results of this body of work. Additional factors contribute to the net power difference between this study and the one using petcoke, but will not be discussed further here. This does not preclude the oxy-combustor option from being economically competitive for other coals, product slates, and/or process constraints.

Table T5-1: Plant power output summary for oxy-combustion case

| Power summary (kW) | Reference case | Sour PSA with oxy-combustor case |
|---|---|---|
| **Gas turbines** | 430,900 | 405,160 |
| **Steam turbine** | 203,800 | 216,694 (186,439+30,255) |
| **Total gross power output** | 634,700 | 621,854 |
| **Total auxiliary power** | 189,410 | 191,521 |
| **Net plant power output** | 445,290 | 430,333 |

Table T5-1 shows a high level breakdown of the power balance of the plant. The higher auxiliary power load compared to the reference case is mainly due a larger ASU required for the higher $O_2$ demand and the power to compress the oxy-combustor exhaust gas that are not fully compensated by the reduction in $CO_2$ compression power or the suppression of the solvent circulating pumps of the Selexol unit. The detail of the different auxiliary loads is provided in table T5-a3 in the Appendix.

### *Acid Gas Enrichment and Claus plant:*

The alternative sulfur disposition 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 to form 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-condensable, including combustibles like $CH_4$ and CO, and is blended with the gas turbine fuel. Figure T5-4 shows a general schematic of this tail gas treatment option.

The implementation of the sour PSA with acid gas enrichment and Claus plant in the coal gasification to power process that we have simulated is described on the simplified PFD provided in figure T5-8.

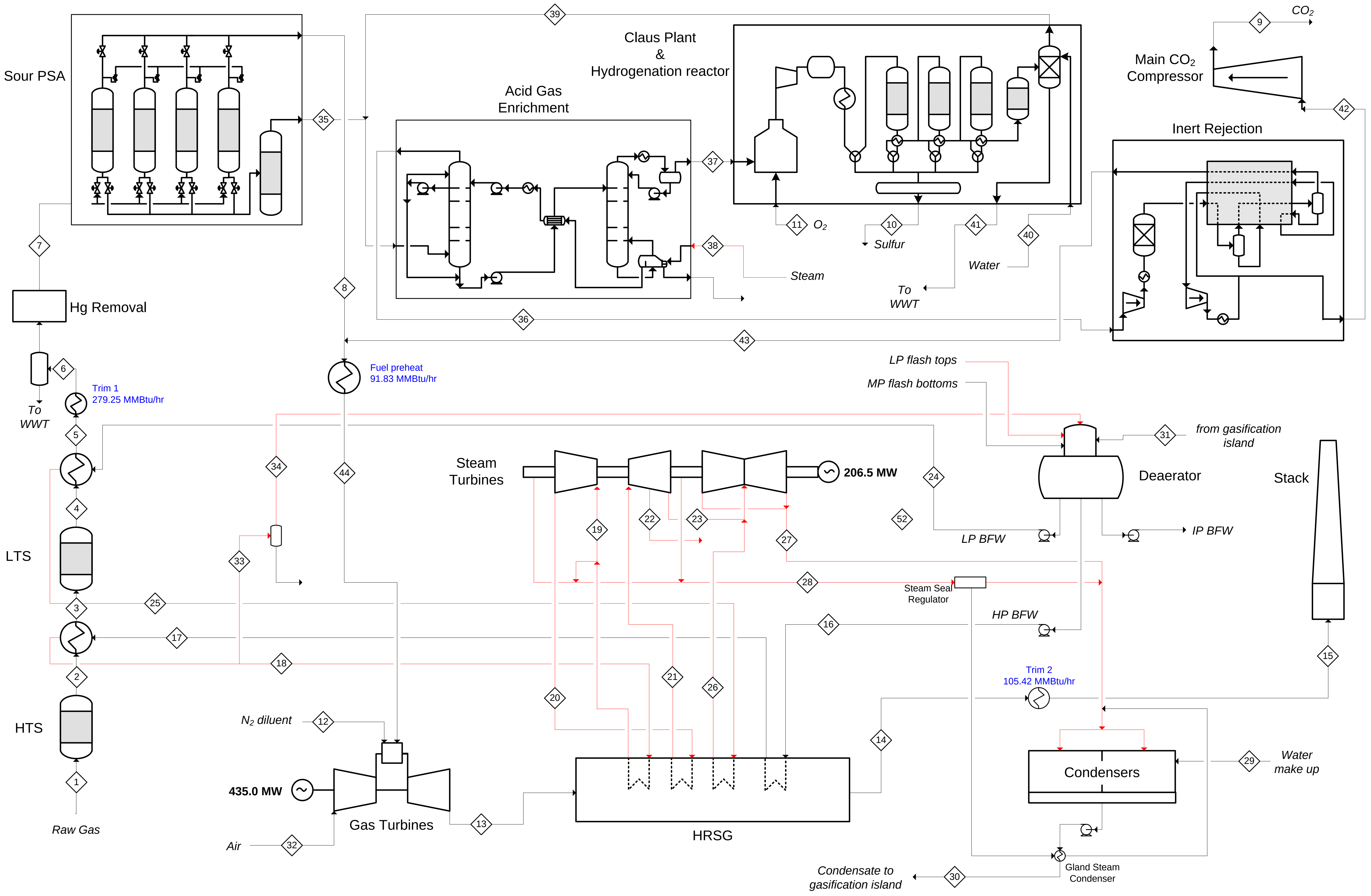


Figure T5-8: Simplified PFD of the Sour PSA with AGE and Claus Plant (adapted from DOE/NETL report 2010/1399 p339)

The AGE system performance was based on Exxon Mobil Flexsorb® technology and consists of an absorber and stripper columns arrangement. We did not model the details of the process but use results from a prior internal study and publically available information to extract correlations and calculate an overall heat and mass balance. The feed stream to the Claus unit has a similar composition to the one of the reference plant therefore the Claus plant is not modeled but instead scaled from the NETL S3B reference plant.

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 which requires 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 that is 446 ppm (v.s. <10 ppm) and $N_2$ level at 4082 ppm (v.s. <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. We take 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 diluting the sour PSA product and the non-condensables from the light impurities rejection unit with $N_2$ in order to achieve a fuel with a LHV of 120 Btu/SCF similar to the reference case. Consequently, there is about 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 turbines power output also translates into more heat being available in the HRSG to produce more steam than the reference case. Again, the approach we take 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 produced is produced at the same conditions and HP, IP and LP ratio than in the reference case. Simulation of the steam cycle indicates that the additional available heat translates into an increase of 1.3% of the steam turbine power output to 206.5 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. Similarly 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 (approximately 92 MMBtu/hr). However, as shown on Figure T5-8, there is still 279 MMBtu/hr available in the raw syngas (Trim1). The estimation of the steam needed for the regeneration of the solvent of the Selexol system in the base case is not readily attainable from the heat and material balance in the NETL report, but we assumed that some of the heat from the raw syngas stream was used for the reboiler duty of the $H_2S$ stripping column. Our estimate of the Flexsorb® process, suggest that the left over heat available in the raw syngas and HRSG is sufficient to generate enough steam for the reboiler to strip the $H_2S$ in the regeneration column of the AGE system.
The summary of the heat and mass balance for the simulated process is available in table T5-a4 in the Appendix.

The choice of the AGE and Claus plant option for sulfur disposal in combination with the Sour PSA technology results in a noticeable increase of the overall power production from both the gas and the steam turbines (1.1% overall increase). 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 higher water circulation and cooling requirement due to the increased steam production.

Table T5-2: Plant power output summary for AGE and Claus case

| Power summary (kW) | Reference case | Sour PSA with AGE + Claus case |
|---|---|---|
| **Gas turbines** | 430,900 | 435015 |
| **Steam turbine** | 203,800 | 206,547 |
| **Total gross power output** | 634,700 | 641,562 |
| **Total auxiliary power** | 189,410 | 191,990 |
| **Net plant power output** | 445,290 | 449,572 |

A more detailed list of the different auxiliary loads is provided in Table T5-a5 in the appendix.

### 3- Economic estimation:

Based on the process simulation results, the cost of the equipment already present in the reference case is estimated with classic scaling methodology, using the same exponent and scaling parameters that NETL used for the reference case.

We have leveraged internal engineering resources to estimate the cost of new equipments such as the sour PSA, $H_2S$ oxy-combustor, AGE, sour compression and inert rejection/$CO_2$ purification units. The cost estimates for the sour PSA, sour compression and inert rejection are based on recent internal information. The $H_2S$ oxy-combustor and AGE cost estimates are based on past studies involving external partners and therefore may carry slightly higher uncertainties.

Since these new technologies have not been implemented at commercial scale, we used a 20% contingency for both the process and the project. This is certainly a conservative approach for the sour PSA systems given the extensive experience that Air Products has with $H_2$ PSA technology at commercial scale. This is particularly of interest in the AGE and Claus plant tail gas treatment case where the technologies employed in the treatment of the tail gas are commercially available. 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 of the Sour PSA with AGE and Claus option compares favorably with the reference case (5.3% cheaper) while the Sour PSA with the oxy-combustion option is slightly more expensive.

### Table T5-3: High level summary of the Total Plant Cost.

| case | Reference S3B | Oxy-Combustor | AGE+Claus |
|---|---|---|---|
| **Total Plant Cost ($x1000)** | 1,573,049 | 1,592,633 | 1,496,764 |
| **Total Plant Cost ($/kW)** | 3,533 | 3,701 | 3,329 |

Table T5-a6 and T5-a7 in the Appendix detail the total plant cost for the sour PSA with oxy-combustion of the tail gas and sour PSA with AGE and Claus plant respectively. Line items in red color indicate departures from the reference case.

Both fixed and variable operating costs have been estimated on the same basis that NETL used for the reference case. The cost of the sour PSA adsorbent is provided for both the initial load and as a yearly cost though it is expected that the adsorbent will be replace only every 10 years. The cost of the natural gas burned in the oxy-combustor was set at $5/MMBtu for the lifetime of the plant.
The NETL report does not include the cost of $CO_2$ transport, sequestration and monitoring (TS&M) in the operating cost but rather add it to the COE. From the NETL a TS&M cost of $6.492 per metric tonne of $CO_2$ was back calculated and used to estimate the annual cost of TS&M for both sour PSA cases. In addition, a variant of the COE calculation was done purely based on cash flow, allowing for TS&M costs to be accounted for in the operating cost.

### Table T5-4: High level summary of the Fixed and Variable Operating Cost.

| case | Reference S3B | Oxy-Combustor | AGE+ Claus |
|---|---|---|---|
| **Power output (kW)** | 445,290 | 430,333 | 449,572 |
| **Fixed Operating Cost ($/year)** | 57,327,773 | 57,942,898 | 54,930,072 |
| **Variable Operating Cost without TS&M ($/year)** | 63,984,538 | 65,182,331 | 65,812,248 |
| **$CO_2$ TS&M Cost ($/year)** | 19,659,715 | 21,048,338 | 19,794,318 |

The complete summary of Operating and variable cost for both Sour PSA option is provided in tables T5-a8 and T5-a9 in the appendix.

In the NETL report, the COE is calculated using a capital charge factor and the Total Overnight Cost (TOC). The TOC is the sum of the Total Plant Cost (see above table T5-3) and the Owner's Cost. We use the same methodology as NETL to calculate Owner's Coast for both sour PSA cases.

### Table T5-5: Summary of the Owner's Cost and Total Overnight Cost.

| Item | | Reference S3B | Oxy-Combustor | AGE+ Claus |
|---|---|---|---|---|
| **Start up cost** | | **48,551** | **49,081** | **46,779** |
| | 6 months operating labor | 12,933 | 13,045 | 12,497 |
| | 1 month maintenance materials at full capacity | 2,861 | 2,897 | 2,722 |
| | 1 month non fuel consumable at full capacity | 199 | 159 | 527 |
| | 1 month waste disposal | 292 | 292 | 292 |
| | 25% of one month's fuel cost at full capacity | 805 | 836 | 805 |
| | 2% of TPC | 31,461 | 31,853 | 29,935 |
| **Inventory Capital** | | **14,797** | **14,830** | **15,065** |
| | 0.5% of TPC | 7,865 | 7,963 | 7,484 |
| | 60 days of supply (full capacity) of fuel (n/a for NG) | 6,352 | 6,352 | 6,352 |
| | 60 days of supply (full capacity) of non fuel consumables. | 579 | 514 | 1,229 |
| **Land** | ($3000/acre) | **900** | **900** | **900** |
| **Financing cost** | (2.7% of TPC) | **42,473** | **43,001** | **40,413** |
| **Other Owner Cost** | (15% of TPC) | **235,959** | **238,895** | **224,515** |
| **Initial Catalyst and Chemical Cost** | | **6,922** | **3,138** | **3,138** |
| **Total Owner's Cost ($x1000)** | | **349,602** | **349,845** | **330,808** |
| **Total Plant Cost ($x1000)** | | **1,573,059** | **1,592,633** | **1,496,764** |
| **Total Overnight Cost ($x1000)** | | **1,922,661** | **1,942,477** | **1,827,432** |

Using the same assumptions, the estimation of the COE suggests a significant advantage for the AGE and Claus plant technology over the oxy-combustion case. The combination of Air Products sour PSA technology with AGE and Claus plant for sulfur disposition also results in a lower COE compared to the reference case (NETL S3B case).

Two separate methods were employed reach the COE for the Sour PSA cases. The first method is to calculate the COE using the same method and assumption that NETL used in their report. The COE is calculated using a capital charge factor of 0.1243 (High risk project) the Total Overnight Cost and Fixed and Variable Operating Costs (without TS&M), the TS&M cost is then added to the calculated COE.. Alternately, the COE can be calculated purely from the cash flow statement. The cash flow statement assumes 12% Internal Rate of Return and uses the same assumption as in the NETL report. These assumptions are listed in table T5-6 for reference.

### Table T5-6: Summary of assumption used for cash flow analysis

| 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 management (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 |

The COE results are represented in Table T5-7 under the row labeled COE ($/MW.h) TS&M on cash flow basis (COE from cash flow analysis), and to under the row labeled COE ($/MW.h) TS&M charge (COE using NETL method). The cash flow summary table for the Sour PSA cases is available in the appendix as table T5-a10 and T5-a11 for the Oxy-combustor and the AGE and Claus plant respectively.

### Table T5-7: cost summary of the studied cases

| Case | Reference S3B | Oxy-Combustor | AGE+ Claus |
|---|---|---|---|
| **Power output (kW)** | 445,290 | 430,333 | 449,572 |
| **Total Overnight Cost ($)** | 1,922,660,634 | 1,942,477,064 | 1,827,572,307 |
| **Fixed Operating Cost ($/year)** | 57,327,773 | 57,942,898 | 54,930,072 |
| **Variable Operating Cost without TS&M ($/year)** | 63,984,538 | 65,182,331 | 65,812,248 |
| **COE ($/MW.h) TS&M charge** | **121.68** | **127.66** | **116.91** |
| **COE($/MW.h) TS&M on cash flow basis** | 123.01 | 129.37 | 118.26 |

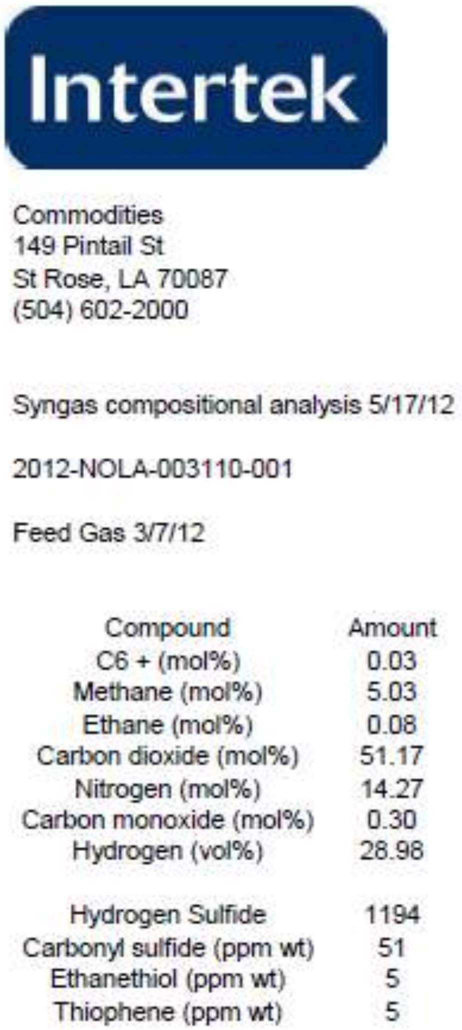
Our economic evaluation shows that the implementation of the sour PSA technology in combination with an AGE and Claus plant for sulfur disposal system could result in up to 3.9% reduction of the cost of electricity produced by IGCC with carbon capture. This corresponds to up to 13.7% reduction of the incremental cost of carbon capture compared to the NETL S3B reference case on a TS&M charge basis.

The COE in the case of the sour PSA with the oxy-combustor option is adversely affected by the constraints we put on the steam cycle that preclude full heat utilization and maximum power production. It is therefore likely that improving heat utilization or, as mentioned earlier, utilizing higher Sulfur containing coals would make the oxy-combustor option more economically competitive.

# Appendix

# Appendix: Task3

## Table T3-a1: Intertek Analysis of Gas Samples from PSA Testing



Intertek

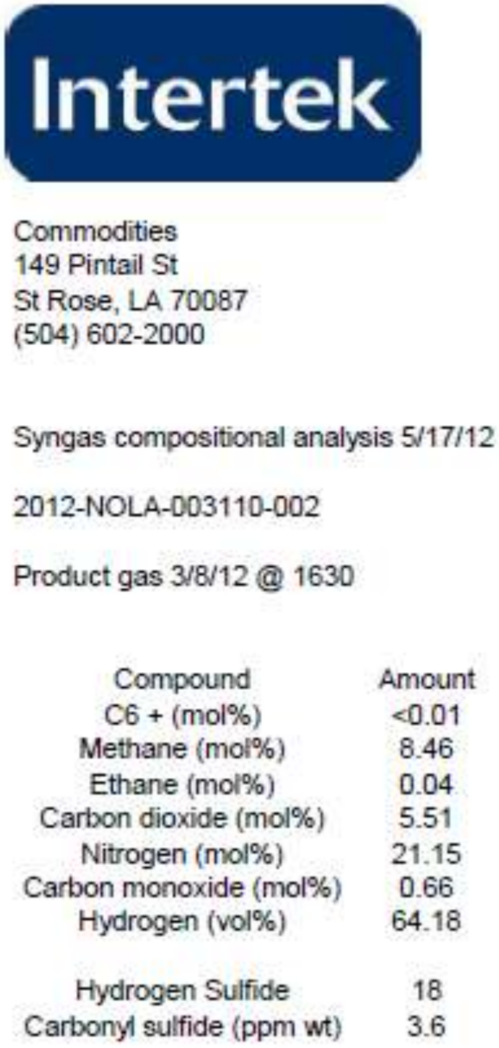
Commodities
149 Pintail St
St Rose, LA 70087
(504) 602-2000

Syngas compositional analysis 5/17/12

2012-NOLA-003110-001

Feed Gas 3/7/12

| Compound | Amount |
|---|---|
| C6 + (mol%) | 0.03 |
| Methane (mol%) | 5.03 |
| Ethane (mol%) | 0.08 |
| Carbon dioxide (mol%) | 51.17 |
| Nitrogen (mol%) | 14.27 |
| Carbon monoxide (mol%) | 0.30 |
| Hydrogen (vol%) | 28.98 |
| Hydrogen Sulfide | 1194 |
| Carbonyl sulfide (ppm wt) | 51 |
| Ethanethiol (ppm wt) | 5 |
| Thiophene (ppm wt) | 5 |



Intertek

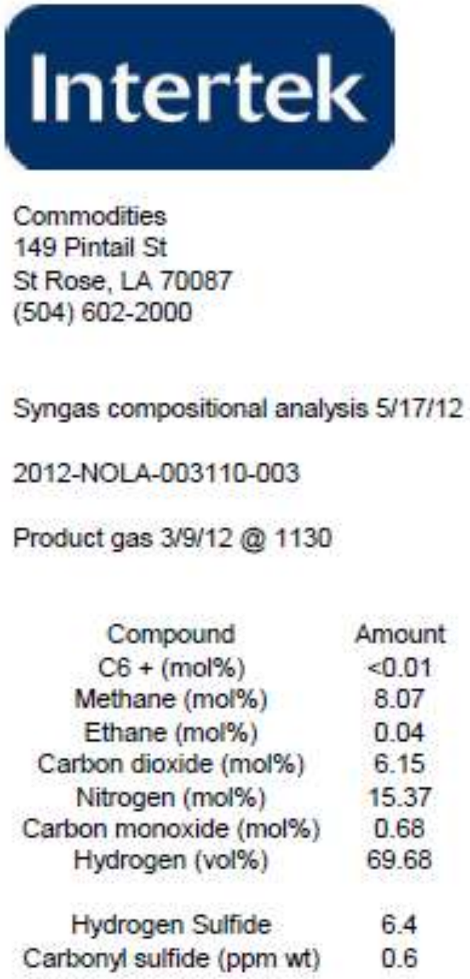
Commodities
149 Pintail St
St Rose, LA 70087
(504) 602-2000

Syngas compositional analysis 5/17/12

2012-NOLA-003110-002

Product gas 3/8/12 @ 1630

| Compound | Amount |
|---|---|
| C6 + (mol%) | <0.01 |
| Methane (mol%) | 8.46 |
| Ethane (mol%) | 0.04 |
| Carbon dioxide (mol%) | 5.51 |
| Nitrogen (mol%) | 21.15 |
| Carbon monoxide (mol%) | 0.66 |
| Hydrogen (vol%) | 64.18 |
| Hydrogen Sulfide | 18 |
| Carbonyl sulfide (ppm wt) | 3.6 |



Intertek

Commodities
149 Pintail St
St Rose, LA 70087
(504) 602-2000

Syngas compositional analysis 5/17/12

2012-NOLA-003110-003

Product gas 3/9/12 @ 1130

| Compound | Amount |
|---|---|
| C6 + (mol%) | <0.01 |
| Methane (mol%) | 8.07 |
| Ethane (mol%) | 0.04 |
| Carbon dioxide (mol%) | 6.15 |
| Nitrogen (mol%) | 15.37 |
| Carbon monoxide (mol%) | 0.68 |
| Hydrogen (vol%) | 69.68 |
| Hydrogen Sulfide | 6.4 |
| Carbonyl sulfide (ppm wt) | 0.6 |

# Appendix: Task5

## Table T5-a1: Heat and Mass Balance for the simulated reference case

| 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 | |

| 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 |

| 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 |

| 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 T5-a2: Heat and Mass Balance for the simulated Sour PSA with Oxy-Combustor case

| stream number | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | | |
| Mole Fraction (%) | | | | | | | | | | |
| $H_2$ | 0.14560 | | | | | | 0.53255 | 0.84158 | 0.00000 | |
| $N_2$ | 0.05240 | | | | | | 0.06631 | 0.10271 | 0.01831 | |
| CO | 0.28300 | | | | | | 0.01010 | 0.01516 | 0.00000 | |
| $CO_2$ | 0.02570 | | | | | | 0.38094 | 0.03063 | 0.95949 | |
| $CH_4$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00000 | |
| Ar | 0.00510 | | | | | | 0.00640 | 0.00991 | 0.00487 | |
| $H_2S$ | 0.00150 | | | | | | 0.00210 | 0.00000 | 0.00000 | |
| COS | 0.00010 | | | | | | 0.00000 | 0.00000 | 0.00000 | |
| $H_2O$ | 0.48540 | | | | | | 0.00160 | 0.00000 | 0.00134 | |
| $SO_2$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00012 | |
| $SO_3$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00000 | |
| $O_2$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.01523 | |
| NO | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00015 | |
| $NO_2$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00000 | |
| $HNO_3$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00048 | |
| $H_2SO_4$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00000 | |
| $H_2SO_3$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00000 | |
| $HNO_2$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00000 | |
| $N_2O_4$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00000 | |
| $C_2H_6$ | 0.00000 | | | | | | 0.00000 | 0.00000 | 0.00000 | |
| | | | | | | | | | | |
| Mole Flow Rate (lbmol/hr) | 77885.00 | | | | | | 61559.68 | 36106.29 | 916.99 | |
| Mass Flow Rate (lb./hr) | 1546091.00 | 1546091.00 | 1546091.00 | 1546091.00 | 1546091.00 | 1546091.00 | 1251807.54 | 243452.68 | 16595.85 | |
| Temperature (F) | 450.00 | 875.30 | 400.00 | 527.40 | 316.00 | 95.00 | 95.00 | 420.00 | 160.53 | |
| Pressure (PSIA) | 579.70 | 569.70 | 564.70 | 552.60 | 552.60 | 552.60 | 522.14 | 464.60 | 2215.00 | |
| Enthalpy (Btu/lb) | -3317.69 | | | | | | -3203.63 | -517.80 | -3789.13 | |

| 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 |
| $N_2$ | | 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 |
| $CO_2$ | | 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 |
| Ar | | 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 |
| COS | | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $H_2O$ | | 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 |
| $SO_3$ | | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $O_2$ | | 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 |
| $NO_2$ | | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $HNO_3$ | | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $H_2SO_4$ | | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $H_2SO_3$ | | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $HNO_2$ | | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $N_2O_4$ | | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $C_2H_6$ | | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| | | | | | | | | | | |
| Mole Flow Rate (lbmol/hr) | | 34759.55 | 263288.19 | 263288.19 | 263288.19 | 37006.78 | 37006.78 | 36200.28 | 42158.47 | 35700.91 |
| Mass Flow Rate (lb./hr) | | 975386.55 | 7236154.51 | 7236154.51 | 7236154.51 | 666677.16 | 666677.16 | 652148.10 | 759484.88 | 643151.89 |
| Temperature (F) | | 385.00 | 1041.90 | 321.87 | 270.00 | 278.80 | 287.16 | 622.26 | 991.91 | 617.74 |
| Pressure (PSIA) | | 384.00 | 15.00 | 15.00 | 15.00 | 2250.70 | 2250.70 | 1814.70 | 1814.70 | 476.70 |
| Enthalpy (Btu/lb) | | 75.05 | -250.42 | -447.15 | -460.72 | -6618.55 | -6610.07 | -6066.10 | -5395.33 | -5559.80 |

| 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 |
| $HNO_3$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $H_2SO_4$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $H_2SO_3$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $HNO_2$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $N_2O_4$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $C_2H_6$ | 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) | 35700.91 | 2205.83 | 39338.22 | 36127.76 | 36127.76 | 36127.76 | 81459.72 | 661.67 | 4034.64 | 86156.03 |
| Mass Flow Rate (lb./hr) | 643151.89 | 39738.00 | 708677.97 | 650841.57 | 650841.57 | 650841.57 | 1467496.81 | 11920.04 | 72684.09 | 1552100.94 |
| Temperature (F) | 1000.00 | 843.13 | 483.47 | 278.80 | 298.02 | 409.00 | 90.06 | 572.22 | 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 | -5913.96 | -5552.42 | -6854.82 | -6810.37 |

| 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 | 0.00000 | 0.00360 | 0.00000 | 0.00000 | 0.00000 |
| $N_2$ | 0.00000 | 0.77590 | 0.00000 | 0.00000 | 0.01467 | 0.01400 | 0.03650 | 0.01400 | 0.99210 | 0.01624 |
| CO | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00292 | 0.00000 | 0.00090 | 0.00000 | 0.00000 | 0.00000 |
| $CO_2$ | 0.00000 | 0.00030 | 0.00000 | 0.00000 | 0.87786 | 0.00000 | 0.00340 | 0.00000 | 0.00000 | 0.85126 |
| $CH_4$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.87000 | 0.00000 | 0.00000 | 0.00000 |
| Ar | 0.00000 | 0.00930 | 0.00000 | 0.00000 | 0.00142 | 0.03600 | 0.00000 | 0.03600 | 0.00230 | 0.00432 |
| $H_2S$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00508 | 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 | 0.00640 | 1.00000 | 1.00000 | 0.00387 | 0.00000 | 0.00000 | 0.00000 | 0.00020 | 0.10576 |
| $SO_2$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00474 |
| $SO_3$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00015 |
| $O_2$ | 0.00000 | 0.20810 | 0.00000 | 0.00000 | 0.00000 | 0.95000 | 0.00070 | 0.95000 | 0.00540 | 0.01645 |
| NO | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00098 |
| $NO_2$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00010 |
| $HNO_3$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $H_2SO_4$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $H_2SO_3$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $HNO_2$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $N_2O_4$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $C_2H_6$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.08460 | 0.00000 | 0.00000 | 0.00000 |
| | | | | | | | | | | |
| Mole Flow Rate (lbmol/hr) | 86156.03 | 208248.64 | 806.50 | 474.68 | 25453.39 | 1946.82 | 85.76 | 218.22 | 37.20 | 26441.13 |
| Mass Flow Rate (lb./hr) | 1552100.94 | 6017315.48 | 14529.06 | 8551.32 | 1008354.85 | 62742.80 | 1521.14 | 7032.89 | 1043.89 | 1080695.55 |
| Temperature (F) | 235.00 | 42.00 | 622.26 | 298.02 | 95.00 | 60.00 | 60.00 | 60.00 | 60.00 | 321.07 |
| Pressure (PSIA) | 110.00 | 13.00 | 1814.70 | 65.00 | 17.64 | 20.00 | 20.00 | 20.00 | 20.00 | 17.64 |
| Enthalpy (Btu/lb) | -6667.31 | -33.35 | -6066.10 | -5690.68 | -3762.74 | -3.82 | -1797.27 | -3.82 | -5.12 | -3754.50 |

| stream number | 41 | 42 | 43 | 44 | 45 | 46 | 47 | 48 | 49 | 50 |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | | |
| Mole Fraction (%) | | | | | | | | | | |
| $H_2$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $N_2$ | 0.01624 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.01831 |
| CO | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $CO_2$ | 0.85126 | 0.00000 | 0.00139 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.95949 |
| $CH_4$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| Ar | 0.00432 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00487 |
| $H_2S$ | 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 |
| $H_2O$ | 0.10576 | 1.00000 | 0.87539 | 1.00000 | 1.00000 | 1.00000 | 1.00000 | 1.00000 | 1.00000 | 0.00134 |
| $SO_2$ | 0.00474 | 0.00000 | 0.00006 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00012 |
| $SO_3$ | 0.00015 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $O_2$ | 0.01645 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.01523 |
| NO | 0.00098 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00015 |
| $NO_2$ | 0.00010 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $HNO_3$ | 0.00000 | 0.00000 | 0.01170 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00048 |
| $H_2SO_4$ | 0.00000 | 0.00000 | 0.11138 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $H_2SO_3$ | 0.00000 | 0.00000 | 0.00002 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $HNO_2$ | 0.00000 | 0.00000 | 0.00006 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $N_2O_4$ | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 | 0.00000 |
| $C_2H_6$ | 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) | 26441.13 | 1102.31 | 1128.47 | 5993.74 | 5993.74 | 6005.43 | 6005.43 | 5922.65 | 5922.65 | 23453.82 |
| Mass Flow Rate (lb./hr) | 1080695.55 | 19858.14 | 31033.35 | 107977.26 | 107977.26 | 108187.91 | 108187.91 | 106696.49 | 106696.49 | 1019958.40 |
| Temperature (F) | 115.00 | 86.00 | 123.61 | 278.00 | 409.00 | 278.00 | 992.00 | 617.74 | 1000.00 | 90.03 |
| Pressure (PSIA) | 17.64 | 435.11 | 435.11 | 65.00 | 65.00 | 2265.70 | 2265.70 | 476.70 | 476.70 | 435.11 |
| Enthalpy (Btu/lb) | -3812.79 | -6819.30 | -5438.30 | -6623.67 | -5633.15 | -6619.33 | -5410.48 | -5559.80 | -5349.08 | -3749.27 |

## Table T5-a3: Power summary for the Sour PSA with Oxy-combustion case

| POWER SUMARY (kW) | reference | Sour PSA w Oxy-Combustor |
|---|---|---|
| Gas Turbine | 430,900 | 405,160 |
| Steam Turbine | 203,800 | 216,694 |
| **Total power** | **634,700** | **621,854** |
| | | |
| **AUXILIARY LOAD (kW)** | | |
| 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 | 73,692 |
| $O_2$ Compressor | 8,670 | 8,670 |
| $N_2$ Compressor | 34,640 | 33,481 |
| $CO_2$ Compressor | 31,220 | 10,890 |
| Boiler Feed Water Pumps | 2,330 | 2,477 |
| Condensate Pump | 220 | 234 |
| Quench Water Pump | 10 | 10 |
| Circulating Water Pump | 3,090 | 3,279 |
| Ground Water Pump | 360 | 383 |
| Cooling Tower Fan | 2,020 | 2,251 |
| Air Cooled Condenser Fan | 2,990 | 3,332 |
| Scrubber Pumps | 750 | 750 |
| Acid Gas Removal/Sour PSA | 18,190 | - |
| CPU | - | 30,963 |
| Gas Turbine Auxiliary | 1,000 | 940 |
| Steam Turbine Auxiliary | 100 | 106 |
| Claus Plant Auxiliary | 250 | - |
| Claus Plant TG Compressor | 1,460 | - |
| Misc. Balance of Plant | 3,000 | 3,000 |
| Transformer Losses | 2,450 | 2,400 |
| **Total Auxiliary Power** | **189,410** | **191,521** |
| | | |
| **NET POWER (kW)** | **445,290** | **430,333** |

## Table T5-a4: Heat and Mass Balance for the simulated Sour PSA AGE and Claus plant case

| 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 | |

| 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 |

| 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 |

| 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 |

| 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 T5-a5: Power summary for the Sour PSA with AGE and Claus plant case

| POWER SUMARY (kW) | reference | Sour PSA w AGE-Claus |
|---|---|---|
| Gas Turbine | 430,900 | 435,015 |
| Steam Turbine | 203,800 | 206,547 |
| **Total power** | **634,700** | **641,562** |
| | | |
| **AUXILIARY LOAD (kW)** | | |
| 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 | 62,762 |
| $O_2$ Compressor | 8,670 | 8,663 |
| $N_2$ Compressor | 34,640 | 35,066 |
| $CO_2$ Compressor | 31,220 | 16,282 |
| Boiler Feed Water Pumps | 2,330 | 2,361 |
| Condensate Pump | 220 | 223 |
| Quench Water Pump | 10 | 10 |
| Circulating Water Pump | 3,090 | 3,131 |
| Ground Water Pump | 360 | 365 |
| Cooling Tower Fan | 2,020 | 2,146 |
| Air Cooled Condenser Fan | 2,990 | 3,177 |
| Scrubber Pumps | 750 | 750 |
| Acid Gas Removal/Sour PSA | 18,190 | - |
| CPU | - | 34,238 |
| Gas Turbine Auxiliary | 1,000 | 1,010 |
| Steam Turbine Auxiliary | 100 | 101 |
| Claus Plant Auxiliary | 250 | 229 |
| Claus Plant TG Compressor | 1,460 | 1,339 |
| Misc. Balance of Plant | 3,000 | 3,000 |
| Transformer Losses | 2,450 | 2,476 |
| **Total Auxiliary Power** | **189,410** | **191,990** |
| | | |
| **NET POWER (kW)** | **445,290** | **449,572** |

## Table T5-a6: Total Plant Cost for the Sour PSA with Oxy-combustor 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** | **$15,910** | **$2,957** | **$12,339** | **$0** | **$0** | **$31,206** | **$2,832** | **0%** | **$0** | **20%** | **$6,808** | **$40,846** | **$95** |
| 1.1 | Coal Receive & Unload | $4,178 | $0 | $2,042 | $0 | $0 | $6,220 | $557 | 0% | $0 | 20% | $1,355 | $8,132 | $19 |
| 1.2 | Coal Stackout & Reclaim | $5,399 | $0 | $1,309 | $0 | $0 | $6,708 | $588 | 0% | $0 | 20% | $1,459 | $8,755 | $20 |
| 1.3 | Coal Conveyors & Yd Crush | $5,020 | $0 | $1,295 | $0 | $0 | $6,315 | $554 | 0% | $0 | 20% | $1,374 | $8,243 | $19 |
| 1.4 | Other Coal Handling | $1,313 | $0 | $300 | $0 | $0 | $1,613 | $141 | 0% | $0 | 20% | $351 | $2,105 | $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 | $2,957 | $7,393 | $0 | $0 | $10,350 | $992 | 0% | $0 | 20% | $2,268 | $13,610 | $32 |
| **2** | **COAL & SORBENT PREP & FEED** | **$119,959** | **$9,995** | **$20,468** | **$0** | **$0** | **$150,422** | **$13,050** | **0%** | **$0** | **20%** | **$32,694** | **$196,166** | **$456** |
| 2.1 | Coal Crushing & Drying | $48,294 | $2,901 | $7,037 | $0 | $0 | $58,232 | $5,025 | 0% | $0 | 20% | $12,651 | $75,908 | $176 |
| 2.2 | Prepared Coal Storage & Feed | $2,076 | $497 | $326 | $0 | $0 | $2,899 | $248 | 0% | $0 | 20% | $629 | $3,776 | $9 |
| 2.3 | Dry Coal Injection System | $68,331 | $793 | $6,346 | $0 | $0 | $75,470 | $6,500 | 0% | $0 | 20% | $16,394 | $98,364 | $229 |
| 2.4 | Misc.Coal Prep & Feed | $1,258 | $915 | $2,745 | $0 | $0 | $4,918 | $452 | 0% | $0 | 20% | $1,074 | $6,444 | $15 |
| 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 | $4,889 | $4,014 | $0 | $0 | $8,903 | $825 | 0% | $0 | 20% | $1,946 | $11,674 | $27 |
| **3** | **FEEDWATER & MISC. BOP SYSTEMS** | **$7,393** | **$5,413** | **$7,602** | **$0** | **$0** | **$20,408** | **$1,931** | **0%** | **$0** | **23%** | **$5,245** | **$27,584** | **$64** |
| 3.1 | Feedwater System | $1,771 | $3,041 | $1,605 | $0 | $0 | $6,417 | $594 | 0% | $0 | 20% | $1,402 | $8,413 | $20 |
| 3.2 | Water Makeup & Pretreating | $553 | $58 | $309 | $0 | $0 | $920 | $88 | 0% | $0 | 30% | $302 | $1,310 | $3 |
| 3.3 | Other Feedwater Subsystems | $969 | $327 | $295 | $0 | $0 | $1,591 | $143 | 0% | $0 | 20% | $347 | $2,081 | $5 |
| 3.4 | Service Water Systems | $316 | $651 | $2,260 | $0 | $0 | $3,227 | $315 | 0% | $0 | 30% | $1,063 | $4,605 | $11 |
| 3.5 | Other Boiler Plant Systems | $1,697 | $657 | $1,629 | $0 | $0 | $3,983 | $378 | 0% | $0 | 20% | $872 | $5,233 | $12 |
| 3.6 | FO Supply Sys & Nat Gas | $286 | $541 | $504 | $0 | $0 | $1,331 | $128 | 0% | $0 | 20% | $292 | $1,751 | $4 |
| 3.7 | Waste Treatment Equipment | $772 | $0 | $471 | $0 | $0 | $1,243 | $121 | 0% | $0 | 30% | $409 | $1,773 | $4 |
| 3.8 | Misc. Power Plant Equipment | $1,029 | $138 | $529 | $0 | $0 | $1,696 | $164 | 0% | $0 | 30% | $558 | $2,418 | $6 |
| **4** | **GASIFIER & ACCESSORIES** | **$348,177** | **$12,820** | **$70,632** | **$0** | **$0** | **$431,629** | **$40,432** | **6%** | **$24,726** | **14%** | **$67,331** | **$564,119** | **$1,311** |
| 4.1 | Gasifier, Syngas Cooler & Auxiliaries (Siemens) | $112,359 | $0 | $52,484 | $0 | $0 | $164,843 | $14,646 | 15% | $24,726 | 15% | $30,632 | $234,848 | $546 |
| 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 | **$207,824** | $0 | w/equip | $0 | $0 | $207,824 | $20,145 | 0% | $0 | 10% | $22,797 | $250,766 | $583 |
| 4.4 | LT Heat Recovery & FG Saturation | $27,994 | $0 | $10,642 | $0 | $0 | $38,636 | $3,771 | 0% | $0 | 20% | $8,481 | $50,888 | $118 |
| 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 | $1,670 | $680 | $0 | $0 | $2,350 | $225 | 0% | $0 | 20% | $515 | $3,090 | $7 |
| 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 | $11,150 | $6,826 | $0 | $0 | $17,976 | $1,645 | 0% | $0 | 25% | $4,905 | $24,527 | $57 |
| **5A** | **GAS CLEANUP & PIPING** | **$107,530** | **$2,119** | **$57,346** | **$0** | **$0** | **$166,995** | **$16,174** | **18%** | **$30,309** | **20%** | **$42,910** | **$256,388** | **$596** |
| 5A.1 | Sour PSA System | **$27,639** | $0 | $23,452 | $0 | $0 | $51,092 | $4,961 | 20% | $10,218 | 20% | $13,254 | $79,525 | $185 |
| 5A.2 | Elemental Sulfur Plant | $0 | $0 | $0 | $0 | $0 | $0 | $0 | 0% | $0 | 20% | $0 | $0 | $0 |
| 5A.3 | Mercury Removal | $1,384 | $0 | $1,054 | $0 | $0 | $2,438 | $235 | 5% | $122 | 20% | $559 | $3,354 | $8 |
| 5A.4 | Shift Reactors | $7,189 | $0 | $2,894 | $0 | $0 | $10,083 | $967 | 0% | $0 | 20% | $2,210 | $13,260 | $31 |
| 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 | $924 | $647 | $0 | $0 | $1,571 | $146 | 0% | $0 | 20% | $343 | $2,060 | $5 |
| 5A.8 | CPU | **$42,908** | $0 | $17,163 | $0 | $0 | $60,071 | $5,827 | 20% | $12,014 | 20% | $15,582 | $93,495 | $217 |
| 5A.8b | Inert Rejection Unit | **$0** | $0 | $0 | $0 | $0 | $0 | $0 | 20% | $0 | 20% | $0 | $0 | $0 |
| 5A9 | Sour Gas Oxycombustor | **$28,410** | $0 | $11,364 | $0 | $0 | $39,774 | $3,858 | 20% | $7,955 | 20% | $10,317 | $61,904 | $144 |
| 5A.10 | HGCU Foundations | $0 | $1,195 | $771 | $0 | $0 | $1,966 | $180 | 0% | $0 | 30% | $644 | $2,790 | $6 |
| **5B** | **CO2 COMPRESSION** | **$7,364** | **$0** | **$4,180** | **$0** | **$0** | **$11,544** | **$1,110** | **0%** | **$0** | **20%** | **$2,531** | **$15,185** | **$35** |
| 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 | **$7,364** | $0 | $4,180 | $0 | $0 | $11,544 | $1,110 | 0% | $0 | 20% | $2,531 | $15,185 | $35 |
| **6** | **COMBUSTION TURBINE/ACCESSORIES** | **$88,688** | **$777** | **$7,217** | **$0** | **$0** | **$96,682** | **$9,163** | **10%** | **$9,505** | **10%** | **$11,893** | **$127,243** | **$296** |
| 6.1 | Combustion Turbine Generator | **$88,688** | $0 | $6,358 | $0 | $0 | $95,046 | $9,010 | 10% | $9,505 | 10% | $11,356 | $124,917 | $290 |
| 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 | $777 | $860 | $0 | $0 | $1,636 | $153 | 0% | $0 | 30% | $537 | $2,327 | $5 |
| **7** | **HRSG, DUCTING & STACK** | **$33,447** | **$2,409** | **$7,460** | **$0** | **$0** | **$43,317** | **$4,097** | **0%** | **$0** | **11%** | **$5,358** | **$52,771** | **$123** |
| 7.1 | Heat Recovery Steam Generator | **$30,069** | $0 | $4,276 | $0 | $0 | $34,345 | $3,265 | 0% | $0 | 10% | $3,761 | $41,371 | $96 |
| 7.2 | Open | $0 | $0 | $0 | $0 | $0 | $0 | $0 | 0% | $0 | 0% | $0 | $0 | $0 |
| 7.3 | Ductwork | $0 | $1,732 | $1,266 | $0 | $0 | $2,998 | $264 | 0% | $0 | 20% | $652 | $3,914 | $9 |
| 7.4 | Stack | **$3,378** | $0 | $1,269 | $0 | $0 | $4,648 | $445 | 0% | $0 | 10% | $509 | $5,602 | $13 |
| 7.9 | HRSG,Duct & Stack Foundations | $0 | $677 | $649 | $0 | $0 | $1,326 | $124 | 0% | $0 | 30% | $435 | $1,885 | $4 |
| **8** | **STEAM TURBINE GENERATOR** | **$57,757** | **$839** | **$14,539** | **$0** | **$0** | **$73,135** | **$7,089** | **0%** | **$0** | **16%** | **$12,979** | **$93,203** | **$217** |
| 8.1 | Steam TG & Accessories | **$24,520** | $0 | $4,089 | $0 | $0 | $28,609 | $2,745 | 0% | $0 | 10% | $3,135 | $34,490 | $80 |
| 8.2 | Turbine Plant Auxiliaries | **$169** | $0 | $388 | $0 | $0 | $557 | $54 | 0% | $0 | 10% | $61 | $673 | $2 |
| 8.3a | Condenser & Auxiliaries | **$2,928** | $0 | $935 | $0 | $0 | $3,863 | $369 | 0% | $0 | 10% | $423 | $4,655 | $11 |
| 8.3b | Air Cooled Condenser | **$26,827** | $0 | $5,378 | $0 | $0 | $32,205 | $3,221 | 0% | $0 | 20% | $7,085 | $42,511 | $99 |
| 8.4 | Steam Piping | **$3,313** | $0 | $2,331 | $0 | $0 | $5,644 | $485 | 0% | $0 | 25% | $1,532 | $7,662 | $18 |
| 8.9 | TG Foundations | $0 | **$839** | $1,418 | $0 | $0 | $2,257 | $214 | 0% | $0 | 30% | $741 | $3,213 | $7 |
| **9** | **COOLING WATER SYSTEM** | **$8,367** | **$7,981** | **$6,832** | **$0** | **$0** | **$23,181** | **$2,154** | **0%** | **$0** | **20%** | **$5,170** | **$30,504** | **$71** |
| 9.1 | Cooling Towers | **$5,829** | $0 | $1,060 | $0 | $0 | $6,889 | $656 | 0% | $0 | 15% | $1,132 | $8,677 | $20 |
| 9.2 | Circulating Water Pumps | **$1,463** | $0 | $94 | $0 | $0 | $1,557 | $131 | 0% | $0 | 15% | $253 | $1,941 | $5 |
| 9.3 | Circ.Water System Auxiliaries | **$127** | $0 | $18 | $0 | $0 | $144 | $14 | 0% | $0 | 15% | $24 | $182 | $0 |
| 9.4 | Circ.Water Piping | $0 | $5,296 | $1,373 | $0 | $0 | $6,669 | $602 | 0% | $0 | 20% | $1,454 | $8,725 | $20 |
| 9.5 | Make-up Water System | **$323** | $0 | $462 | $0 | $0 | $784 | $76 | 0% | $0 | 20% | $172 | $1,032 | $2 |
| 9.6 | Component Cooling Water Sys | **$626** | $749 | $533 | $0 | $0 | $1,908 | $179 | 0% | $0 | 20% | $417 | $2,504 | $6 |
| 9.9 | Circ.Water System Foundations | $0 | **$1,937** | $3,293 | $0 | $0 | $5,229 | $496 | 0% | $0 | 30% | $1,718 | $7,443 | $17 |
| **10** | **ASH/SPENT SORBENT HANDLING SYS** | **$21,399** | **$1,473** | **$10,611** | **$0** | **$0** | **$33,483** | **$3,213** | **0%** | **$0** | **11%** | **$3,974** | **$40,670** | **$95** |
| 10.1 | Slag Dewatering & Cooling | $18,925 | $0 | $9,333 | $0 | $0 | $28,258 | $2,715 | 0% | $0 | 10% | $3,097 | $34,070 | $79 |
| 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 | $561 | $0 | $610 | $0 | $0 | $1,171 | $114 | 0% | $0 | 15% | $193 | $1,478 | $3 |
| 10.7 | Ash Transport & Feed Equipment | $752 | $0 | $181 | $0 | $0 | $933 | $87 | 0% | $0 | 15% | $153 | $1,173 | $3 |
| 10.8 | Misc. Ash Handling Equipment | $1,161 | $1,423 | $425 | $0 | $0 | $3,009 | $286 | 0% | $0 | 15% | $494 | $3,789 | $9 |
| 10.9 | Ash/Spent Sorbent Foundation | $0 | $50 | $62 | $0 | $0 | $112 | $11 | 0% | $0 | 30% | $37 | $160 | $0 |
| **11** | **ACCESSORY ELECTRIC PLANT** | **$29,961** | **$12,389** | **$24,016** | **$0** | **$0** | **$66,366** | **$5,718** | **0%** | **$0** | **19%** | **$13,763** | **$85,848** | **$199** |
| 11.1 | Generator Equipment | **$902** | $0 | $893 | $0 | $0 | $1,795 | $171 | 0% | $0 | 10% | $197 | $2,162 | $5 |
| 11.2 | Station Service Equipment | **$4,656** | $0 | $420 | $0 | $0 | $5,076 | $468 | 0% | $0 | 10% | $554 | $6,099 | $14 |
| 11.3 | Switchgear & Motor Control | **$8,609** | $0 | $1,565 | $0 | $0 | $10,174 | $943 | 0% | $0 | 15% | $1,668 | $12,786 | $30 |
| 11.4 | Conduit & Cable Tray | $0 | **$3,999** | $13,193 | $0 | $0 | $17,192 | $1,663 | 0% | $0 | 25% | $4,714 | $23,568 | $55 |
| 11.5 | Wire & Cable | $0 | **$7,640** | $5,020 | $0 | $0 | $12,660 | $919 | 0% | $0 | 25% | $3,395 | $16,974 | $39 |
| 11.6 | Protective Equipment | $0 | **$611** | $2,225 | $0 | $0 | $2,836 | $277 | 0% | $0 | 15% | $467 | $3,580 | $8 |
| 11.7 | Standby Equipment | **$216** | $0 | $211 | $0 | $0 | $427 | $41 | 0% | $0 | 15% | $70 | $537 | $1 |
| 11.8 | Main Power Transformers | **$15,578** | $0 | $123 | $0 | $0 | $15,701 | $1,187 | 0% | $0 | 15% | $2,533 | $19,422 | $45 |
| 11.9 | Electrical Foundations | $0 | **$139** | $366 | $0 | $0 | $505 | $48 | 0% | $0 | 30% | $166 | $719 | $2 |
| **12** | **INSTRUMENTATION & CONTROL** | **$10,765** | **$1,981** | **$6,936** | **$0** | **$0** | **$19,682** | **$1,785** | **5%** | **$984** | **17%** | **$3,741** | **$26,192** | **$61** |
| 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 | **$1,063** | $0 | $710 | $0 | $0 | $1,773 | $168 | 5% | $89 | 15% | $304 | $2,334 | $5 |
| 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 | **$244** | $0 | $156 | $0 | $0 | $401 | $38 | 5% | $20 | 20% | $92 | $550 | $1 |
| 12.7 | Computer & Accessories | **$5,669** | $0 | $181 | $0 | $0 | $5,850 | $537 | 5% | $293 | 10% | $668 | $7,348 | $17 |
| 12.8 | Instrument Wiring & Tubing | $0 | **$1,981** | $4,049 | $0 | $0 | $6,030 | $512 | 5% | $301 | 25% | $1,711 | $8,554 | $20 |
| 12.9 | Other I & C Equipment | **$3,789** | $0 | $1,840 | $0 | $0 | $5,629 | $530 | 5% | $281 | 15% | $966 | $7,406 | $17 |
| **13** | **IMPROVEMENTS TO SITE** | **$3,256** | **$1,919** | **$8,032** | **$0** | **$0** | **$13,207** | **$1,304** | **0%** | **$0** | **30%** | **$4,353** | **$18,864** | **$44** |
| 13.1 | Site Preparation | $0 | **$102** | $2,183 | $0 | $0 | $2,285 | $227 | 0% | $0 | 30% | $754 | $3,266 | $8 |
| 13.2 | Site Improvements | $0 | **$1,817** | $2,414 | $0 | $0 | $4,231 | $417 | 0% | $0 | 30% | $1,394 | $6,043 | $14 |
| 13.3 | Site Facilities | **$3,256** | $0 | $3,435 | $0 | $0 | $6,691 | $660 | 0% | $0 | 30% | $2,205 | $9,556 | $22 |
| **14** | **BUILDINGS & STRUCTURES** | **$0** | **$6,312** | **$7,108** | **$0** | **$0** | **$13,421** | **$1,220** | **0%** | **$0** | **16%** | **$2,408** | **$17,049** | **$40** |
| 14.1 | Combustion Turbine Area | $0 | **$265** | $150 | $0 | $0 | $415 | $36 | 0% | $0 | 20% | $90 | $541 | $1 |
| 14.2 | Steam Turbine Building | $0 | **$2,042** | $2,908 | $0 | $0 | $4,950 | $456 | 0% | $0 | 15% | $811 | $6,217 | $14 |
| 14.3 | Administration Building | $0 | **$860** | $624 | $0 | $0 | $1,484 | $132 | 0% | $0 | 15% | $242 | $1,858 | $4 |
| 14.4 | Circulation Water Pumphouse | $0 | **$162** | $86 | $0 | $0 | $248 | $22 | 0% | $0 | 15% | $40 | $310 | $1 |
| 14.5 | Water Treatment Buildings | $0 | **$482** | $470 | $0 | $0 | $952 | $86 | 0% | $0 | 15% | $156 | $1,194 | $3 |
| 14.6 | Machine Shop | $0 | **$440** | $301 | $0 | $0 | $742 | $66 | 0% | $0 | 15% | $121 | $929 | $2 |
| 14.7 | Warehouse | $0 | **$711** | $459 | $0 | $0 | $1,169 | $103 | 0% | $0 | 15% | $191 | $1,463 | $3 |
| 14.8 | Other Buildings & Structures | $0 | **$416** | $324 | $0 | $0 | $741 | $66 | 0% | $0 | 20% | $161 | $968 | $2 |
| 14.9 | Waste Treating Building & Str. | $0 | **$934** | $1,785 | $0 | $0 | $2,719 | $254 | 0% | $0 | 20% | $595 | $3,568 | $8 |
| | **TOTAL COST** | **$859,975** | **$69,384** | **$265,319** | **$0** | **$0** | **$1,194,678** | **$111,272** | **5%** | **$65,524** | **16%** | **$221,158** | **$1,592,633** | **$3,701** |

## Table T5-a7: Total Plant Cost for the Sour PSA with AGE and Claus plant 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** | **$15,910** | **$2,957** | **$12,339** | **$0** | **$0** | **$31,206** | **$2,832** | **0%** | **$0** | **20%** | **$6,808** | **$40,846** | **$91** |
| 1.1 | Coal Receive & Unload | $4,178 | $0 | $2,042 | $0 | $0 | $6,220 | $557 | 0% | $0 | 20% | $1,355 | $8,132 | $18 |
| 1.2 | Coal Stackout & Reclaim | $5,399 | $0 | $1,309 | $0 | $0 | $6,708 | $588 | 0% | $0 | 20% | $1,459 | $8,755 | $19 |
| 1.3 | Coal Conveyors & Yd Crush | $5,020 | $0 | $1,295 | $0 | $0 | $6,315 | $554 | 0% | $0 | 20% | $1,374 | $8,243 | $18 |
| 1.4 | Other Coal Handling | $1,313 | $0 | $300 | $0 | $0 | $1,613 | $141 | 0% | $0 | 20% | $351 | $2,105 | $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 | $2,957 | $7,393 | $0 | $0 | $10,350 | $992 | 0% | $0 | 20% | $2,268 | $13,610 | $30 |
| **2** | **COAL & SORBENT PREP & FEED** | **$119,959** | **$9,995** | **$20,468** | **$0** | **$0** | **$150,422** | **$13,050** | **0%** | **$0** | **20%** | **$32,694** | **$196,166** | **$436** |
| 2.1 | Coal Crushing & Drying | $48,294 | $2,901 | $7,037 | $0 | $0 | $58,232 | $5,025 | 0% | $0 | 20% | $12,651 | $75,908 | $169 |
| 2.2 | Prepared Coal Storage & Feed | $2,076 | $497 | $326 | $0 | $0 | $2,899 | $248 | 0% | $0 | 20% | $629 | $3,776 | $8 |
| 2.3 | Dry Coal Injection System | $68,331 | $793 | $6,346 | $0 | $0 | $75,470 | $6,500 | 0% | $0 | 20% | $16,394 | $98,364 | $219 |
| 2.4 | Misc.Coal Prep & Feed | $1,258 | $915 | $2,745 | $0 | $0 | $4,918 | $452 | 0% | $0 | 20% | $1,074 | $6,444 | $14 |
| 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 | $4,889 | $4,014 | $0 | $0 | $8,903 | $825 | 0% | $0 | 20% | $1,946 | $11,674 | $26 |
| **3** | **FEEDWATER & MISC. BOP SYSTEMS** | **$7,393** | **$5,413** | **$7,602** | **$0** | **$0** | **$20,408** | **$1,931** | **0%** | **$0** | **23%** | **$5,245** | **$27,584** | **$61** |
| 3.1 | Feedwater System | $1,771 | $3,041 | $1,605 | $0 | $0 | $6,417 | $594 | 0% | $0 | 20% | $1,402 | $8,413 | $19 |
| 3.2 | Water Makeup & Pretreating | $553 | $58 | $309 | $0 | $0 | $920 | $88 | 0% | $0 | 30% | $302 | $1,310 | $3 |
| 3.3 | Other Feedwater Subsystems | $969 | $327 | $295 | $0 | $0 | $1,591 | $143 | 0% | $0 | 20% | $347 | $2,081 | $5 |
| 3.4 | Service Water Systems | $316 | $651 | $2,260 | $0 | $0 | $3,227 | $315 | 0% | $0 | 30% | $1,063 | $4,605 | $10 |
| 3.5 | Other Boiler Plant Systems | $1,697 | $657 | $1,629 | $0 | $0 | $3,983 | $378 | 0% | $0 | 20% | $872 | $5,233 | $12 |
| 3.6 | FO Supply Sys & Nat Gas | $286 | $541 | $504 | $0 | $0 | $1,331 | $128 | 0% | $0 | 20% | $292 | $1,751 | $4 |
| 3.7 | Waste Treatment Equipment | $772 | $0 | $471 | $0 | $0 | $1,243 | $121 | 0% | $0 | 30% | $409 | $1,773 | $4 |
| 3.8 | Misc. Power Plant Equipment | $1,029 | $138 | $529 | $0 | $0 | $1,696 | $164 | 0% | $0 | 30% | $558 | $2,418 | $5 |
| **4** | **GASIFIER & ACCESSORIES** | **$326,086** | **$12,820** | **$70,199** | **$0** | **$0** | **$409,105** | **$38,251** | **6%** | **$24,726** | **14%** | **$64,790** | **$536,872** | **$1,194** |
| 4.1 | Gasifier, Syngas Cooler & Auxiliaries (Siemens) | $112,359 | $0 | $52,484 | $0 | $0 | $164,843 | $14,646 | 15% | $24,726 | 15% | $30,632 | $234,848 | $522 |
| 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 | **$185,733** | $0 | w/equip | $0 | $0 | $185,733 | $18,003 | 0% | $0 | 10% | $20,374 | $224,110 | $498 |
| 4.4 | LT Heat Recovery & FG Saturation | $27,994 | $0 | $10,642 | $0 | $0 | $38,636 | $3,771 | 0% | $0 | 20% | $8,481 | $50,888 | $113 |
| 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 | $1,670 | $680 | $0 | $0 | $2,350 | $225 | 0% | $0 | 20% | $515 | $3,090 | $7 |
| 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 | $11,150 | $6,393 | $0 | $0 | $17,543 | $1,606 | 0% | $0 | 25% | $4,787 | $23,936 | $53 |
| **5A** | **GAS CLEANUP & PIPING** | **$69,482** | **$2,649** | **$46,111** | **$0** | **$0** | **$118,242** | **$11,450** | **15%** | **$18,317** | **20%** | **$29,741** | **$177,750** | **$395** |
| 5A.1 | Sour PSA System | **$27,639** | $0 | $23,452 | $0 | $0 | $51,092 | $4,961 | 20% | $10,218 | 20% | $13,254 | $79,525 | $177 |
| 5A.2 | Elemental Sulfur Plant | $4,782 | $953 | $6,170 | $0 | $0 | $11,905 | $1,156 | 0% | $0 | 20% | $2,612 | $15,674 | $35 |
| 5A.3 | Mercury Removal | $1,384 | $0 | $1,054 | $0 | $0 | $2,438 | $235 | 5% | $122 | 20% | $559 | $3,354 | $7 |
| 5A.4 | Shift Reactors | $7,189 | $0 | $2,894 | $0 | $0 | $10,083 | $967 | 0% | $0 | 20% | $2,210 | $13,260 | $29 |
| 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 | $924 | $647 | $0 | $0 | $1,571 | $146 | 0% | $0 | 20% | $343 | $2,060 | $5 |
| 5A.8 | AGE | **$6,488** | $0 | $2,595 | $0 | $0 | $9,083 | $881 | 20% | $1,817 | 20% | $2,356 | $14,137 | $31 |
| 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 | $772 | $498 | $0 | $0 | $1,270 | $116 | 0% | $0 | 30% | $416 | $1,803 | $4 |
| **5B** | **CO2 COMPRESSION** | **$10,491** | **$0** | **$5,955** | **$0** | **$0** | **$16,447** | **$1,581** | **0%** | **$0** | **20%** | **$3,606** | **$21,633** | **$48** |
| 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 | **$10,491** | $0 | $5,955 | $0 | $0 | $16,447 | $1,581 | 0% | $0 | 20% | $3,606 | $21,633 | $48 |
| **6** | **COMBUSTION TURBINE/ACCESSORIES** | **$92,553** | **$811** | **$7,255** | **$0** | **$0** | **$100,619** | **$9,536** | **10%** | **$9,891** | **10%** | **$12,378** | **$132,425** | **$295** |
| 6.1 | Combustion Turbine Generator | **$92,553** | $0 | $6,358 | $0 | $0 | $98,911 | $9,377 | 10% | $9,891 | 10% | $11,818 | $129,997 | $289 |
| 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 | $811 | $897 | $0 | $0 | $1,708 | $160 | 0% | $0 | 30% | $560 | $2,428 | $5 |
| **7** | **HRSG, DUCTING & STACK** | **$35,082** | **$2,526** | **$7,824** | **$0** | **$0** | **$45,431** | **$4,297** | **0%** | **$0** | **11%** | **$5,619** | **$55,348** | **$123** |
| 7.1 | Heat Recovery Steam Generator | **$31,540** | $0 | $4,485 | $0 | $0 | $36,025 | $3,425 | 0% | $0 | 10% | $3,945 | $43,395 | $97 |
| 7.2 | Open | $0 | $0 | $0 | $0 | $0 | $0 | $0 | 0% | $0 | 0% | $0 | $0 | $0 |
| 7.3 | Ductwork | $0 | $1,816 | $1,328 | $0 | $0 | $3,144 | $276 | 0% | $0 | 20% | $684 | $4,104 | $9 |
| 7.4 | Stack | **$3,542** | $0 | $1,331 | $0 | $0 | $4,872 | $466 | 0% | $0 | 10% | $534 | $5,872 | $13 |
| 7.9 | HRSG,Duct & Stack Foundations | $0 | $710 | $681 | $0 | $0 | $1,391 | $130 | 0% | $0 | 30% | $456 | $1,976 | $4 |
| **8** | **STEAM TURBINE GENERATOR** | **$55,840** | **$810** | **$14,053** | **$0** | **$0** | **$70,704** | **$6,854** | **0%** | **$0** | **16%** | **$12,547** | **$90,104** | **$200** |
| 8.1 | Steam TG & Accessories | **$23,699** | $0 | $3,952 | $0 | $0 | $27,652 | $2,653 | 0% | $0 | 10% | $3,030 | $33,335 | $74 |
| 8.2 | Turbine Plant Auxiliaries | **$164** | $0 | $375 | $0 | $0 | $538 | $53 | 0% | $0 | 10% | $59 | $650 | $1 |
| 8.3a | Condenser & Auxiliaries | **$2,831** | $0 | $904 | $0 | $0 | $3,736 | $357 | 0% | $0 | 10% | $409 | $4,503 | $10 |
| 8.3b | Air Cooled Condenser | **$25,946** | $0 | $5,202 | $0 | $0 | $31,148 | $3,115 | 0% | $0 | 20% | $6,853 | $41,115 | $91 |
| 8.4 | Steam Piping | **$3,200** | $0 | $2,251 | $0 | $0 | $5,451 | $469 | 0% | $0 | 25% | $1,480 | $7,399 | $16 |
| 8.9 | TG Foundations | $0 | **$810** | $1,369 | $0 | $0 | $2,179 | $207 | 0% | $0 | 30% | $716 | $3,102 | $7 |
| **9** | **COOLING WATER SYSTEM** | **$8,099** | **$7,759** | **$6,638** | **$0** | **$0** | **$22,496** | **$2,090** | **0%** | **$0** | **20%** | **$5,020** | **$29,606** | **$66** |
| 9.1 | Cooling Towers | **$5,637** | $0 | $1,026 | $0 | $0 | $6,663 | $634 | 0% | $0 | 15% | $1,095 | $8,392 | $19 |
| 9.2 | Circulating Water Pumps | **$1,418** | $0 | $91 | $0 | $0 | $1,508 | $127 | 0% | $0 | 15% | $245 | $1,881 | $4 |
| 9.3 | Circ.Water System Auxiliaries | **$123** | $0 | $17 | $0 | $0 | $140 | $13 | 0% | $0 | 15% | $23 | $176 | $0 |
| 9.4 | Circ.Water Piping | $0 | $5,148 | $1,335 | $0 | $0 | $6,482 | $586 | 0% | $0 | 20% | $1,414 | $8,482 | $19 |
| 9.5 | Make-up Water System | **$314** | $0 | $449 | $0 | $0 | $762 | $74 | 0% | $0 | 20% | $167 | $1,003 | $2 |
| 9.6 | Component Cooling Water Sys | **$608** | $727 | $517 | $0 | $0 | $1,852 | $173 | 0% | $0 | 20% | $405 | $2,431 | $5 |
| 9.9 | Circ.Water System Foundations | $0 | **$1,884** | $3,204 | $0 | $0 | $5,088 | $483 | 0% | $0 | 30% | $1,671 | $7,242 | $16 |
| **10** | **ASH/SPENT SORBENT HANDLING SYS** | **$21,399** | **$1,473** | **$10,611** | **$0** | **$0** | **$33,483** | **$3,213** | **0%** | **$0** | **11%** | **$3,974** | **$40,670** | **$90** |
| 10.1 | Slag Dewatering & Cooling | $18,925 | $0 | $9,333 | $0 | $0 | $28,258 | $2,715 | 0% | $0 | 10% | $3,097 | $34,070 | $76 |
| 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 | $561 | $0 | $610 | $0 | $0 | $1,171 | $114 | 0% | $0 | 15% | $193 | $1,478 | $3 |
| 10.7 | Ash Transport & Feed Equipment | $752 | $0 | $181 | $0 | $0 | $933 | $87 | 0% | $0 | 15% | $153 | $1,173 | $3 |
| 10.8 | Misc. Ash Handling Equipment | $1,161 | $1,423 | $425 | $0 | $0 | $3,009 | $286 | 0% | $0 | 15% | $494 | $3,789 | $8 |
| 10.9 | Ash/Spent Sorbent Foundation | $0 | $50 | $62 | $0 | $0 | $112 | $11 | 0% | $0 | 30% | $37 | $160 | $0 |
| **11** | **ACCESSORY ELECTRIC PLANT** | **$29,953** | **$12,405** | **$24,023** | **$0** | **$0** | **$66,381** | **$5,719** | **0%** | **$0** | **19%** | **$13,774** | **$85,874** | **$191** |
| 11.1 | Generator Equipment | **$877** | $0 | $868 | $0 | $0 | $1,745 | $166 | 0% | $0 | 10% | $191 | $2,102 | $5 |
| 11.2 | Station Service Equipment | **$4,661** | $0 | $420 | $0 | $0 | $5,081 | $469 | 0% | $0 | 10% | $555 | $6,105 | $14 |
| 11.3 | Switchgear & Motor Contol | **$8,618** | $0 | $1,567 | $0 | $0 | $10,185 | $944 | 0% | $0 | 15% | $1,669 | $12,799 | $28 |
| 11.4 | Conduit & Cable Tray | $0 | **$4,003** | $13,207 | $0 | $0 | $17,210 | $1,665 | 0% | $0 | 25% | $4,719 | $23,593 | $52 |
| 11.5 | Wire & Cable | $0 | **$7,648** | $5,025 | $0 | $0 | $12,674 | $920 | 0% | $0 | 25% | $3,398 | $16,992 | $38 |
| 11.6 | Protective Equipment | $0 | **$611** | $2,225 | $0 | $0 | $2,836 | $277 | 0% | $0 | 15% | $467 | $3,580 | $8 |
| 11.7 | Standby Equipment | **$219** | $0 | $214 | $0 | $0 | $433 | $41 | 0% | $0 | 15% | $71 | $546 | $1 |
| 11.8 | Main Power Transformers | **$15,578** | $0 | $123 | $0 | $0 | $15,701 | $1,187 | 0% | $0 | 15% | $2,533 | $19,422 | $43 |
| 11.9 | Electrical Foundations | $0 | **$142** | $374 | $0 | $0 | $516 | $49 | 0% | $0 | 30% | $170 | $735 | $2 |
| **12** | **INSTRUMENTATION & CONTROL** | **$10,769** | **$1,981** | **$6,938** | **$0** | **$0** | **$19,689** | **$1,785** | **5%** | **$984** | **17%** | **$3,742** | **$26,200** | **$58** |
| 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 | **$1,063** | $0 | $710 | $0 | $0 | $1,773 | $168 | 5% | $89 | 15% | $305 | $2,335 | $5 |
| 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 | **$244** | $0 | $156 | $0 | $0 | $401 | $38 | 5% | $20 | 20% | $92 | $551 | $1 |
| 12.7 | Computer & Accessories | **$5,671** | $0 | $181 | $0 | $0 | $5,852 | $537 | 5% | $293 | 10% | $668 | $7,350 | $16 |
| 12.8 | Instrument Wiring & Tubing | $0 | **$1,981** | $4,050 | $0 | $0 | $6,032 | $512 | 5% | $302 | 25% | $1,711 | $8,556 | $19 |
| 12.9 | Other I & C Equipment | **$3,791** | $0 | $1,840 | $0 | $0 | $5,631 | $530 | 5% | $282 | 15% | $966 | $7,409 | $16 |
| **13** | **IMPROVEMENTS TO SITE** | **$3,241** | **$1,910** | **$7,996** | **$0** | **$0** | **$13,148** | **$1,298** | **0%** | **$0** | **30%** | **$4,334** | **$18,780** | **$42** |
| 13.1 | Site Preparation | $0 | **$102** | $2,173 | $0 | $0 | $2,275 | $226 | 0% | $0 | 30% | $750 | $3,251 | $7 |
| 13.2 | Site Improvements | $0 | **$1,809** | $2,403 | $0 | $0 | $4,212 | $416 | 0% | $0 | 30% | $1,388 | $6,016 | $13 |
| 13.3 | Site Facilities | **$3,241** | $0 | $3,420 | $0 | $0 | $6,661 | $657 | 0% | $0 | 30% | $2,195 | $9,513 | $21 |
| **14** | **BUILDINGS & STRUCTURES** | **$0** | **$6,258** | **$7,049** | **$0** | **$0** | **$13,306** | **$1,210** | **0%** | **$0** | **16%** | **$2,388** | **$16,905** | **$38** |
| 14.1 | Combustion Turbine Area | $0 | **$265** | $150 | $0 | $0 | $415 | $36 | 0% | $0 | 20% | $90 | $541 | $1 |
| 14.2 | Steam Turbine Building | $0 | **$2,023** | $2,881 | $0 | $0 | $4,904 | $452 | 0% | $0 | 15% | $803 | $6,158 | $14 |
| 14.3 | Administration Building | $0 | **$855** | $620 | $0 | $0 | $1,475 | $131 | 0% | $0 | 15% | $241 | $1,848 | $4 |
| 14.4 | Circulation Water Pumphouse | $0 | **$159** | $84 | $0 | $0 | $243 | $21 | 0% | $0 | 15% | $40 | $304 | $1 |
| 14.5 | Water Treatment Buildings | $0 | **$466** | $455 | $0 | $0 | $921 | $83 | 0% | $0 | 15% | $151 | $1,155 | $3 |
| 14.6 | Machine Shop | $0 | **$438** | $300 | $0 | $0 | $738 | $66 | 0% | $0 | 15% | $121 | $924 | $2 |
| 14.7 | Warehouse | $0 | **$707** | $456 | $0 | $0 | $1,163 | $103 | 0% | $0 | 15% | $190 | $1,455 | $3 |
| 14.8 | Other Buildings & Structures | $0 | **$414** | $323 | $0 | $0 | $737 | $66 | 0% | $0 | 20% | $160 | $963 | $2 |
| 14.9 | Waste Treating Building & Str. | $0 | **$931** | $1,780 | $0 | $0 | $2,710 | $253 | 0% | $0 | 20% | $593 | $3,556 | $8 |
| | **TOTAL COST** | **$806,259** | **$69,767** | **$255,062** | **$0** | **$0** | **$1,131,087** | **$105,098** | **5%** | **$53,919** | **16%** | **$206,659** | **$1,496,764** | **$3,329** |

## Table T5-a8: Operating Cost for the Sour PSA with Oxy-combustor case

**Operating and Maintenance Labor**

| | | | | |
|---|---|---|---|---|
| **Operating Labor** | | | | |
| Operating Labor Rate | 34.65 | $/hour | | |
| Operating Labor Burden Rate | 30% | of base | | |
| Labor Over Head Charge Rate | 25% | of labor | | |
| Administrative & Support Labor | 25% | of burdened O&M labor | | |
| Maintenance labor | 0.914% | of TPC* | | |
| | | | | |
| **Work force** | **16** | | | |
| Skilled Operator | 2 | | | |
| Operator | 10 | | | |
| Foreman | 1 | | | |
| Lab Technician, etc | 3 | | | |
| | | | **Annual Cost ($)** | **Annual Unit Cost ($/kW)** |
| **Fixed Operating Costs** | | | | |
| Operating Labor | | | $6,313,507 | $14,671.204 |
| Maintenance Labor | | | $14,558,691 | $33,831.200 |
| Administrative & Support Labor | | | $5,218,050 | $12,125.601 |
| Property Taxes & Insurance | | | $31,852,651 | $74,018.563 |
| | | | | |
| **TOTAL FIXED OPERATING COST** | | | **$57,942,898** | **$134,646.567** |

* (back calculated from NETL)

**Variable Operating Costs**

| | | | | **Annual Cost ($)** | **Annual Unit Cost ($/kW.h)** |
|---|---|---|---|---|---|
| **Maintenance Material Cost** | 1.745% | of TPC* | | **$ 27,790,979** | **$0.00922** |

| | **Consumption** | | | | | |
|---|---|---|---|---|---|---|
| | **Initial** | **per day** | **Unit Cost ($)** | **Initial Cost ($)** | **Annual Cost ($)** | **Annual Unit Cost ($/kW.h)** |
| **Water (1000 gallons)** | 0.00 | 3,097 | 1.08 | **$0** | **$978,081** | **$0.00032** |
| | | | | | | |
| **Chemicals** | | | | **$3,137,874** | **$1,524,143** | **$0.00051** |
| MU & WT Chem (lb) | 0 | 17,265 | 0.17 | $0 | $872,524 | $0.00029 |
| Carbon (Mercury Removal) (lb) | 114,477 | 157 | 1.05 | $120,220 | $48,088 | $0.00002 |
| COS Catalyst (m3) | 0 | 0 | 2397.36 | $0 | $0 | $0.00000 |
| Water Gas Shift Catalyst (ft3) | 6,049 | 4.14 | 498.83 | $3,017,654 | $603,531 | $0.00020 |
| AGE (Flexorb) Solution (Gal) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| 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 | $0 | $0 | $0.00000 |
| Claus Catalyst (ft3) | 0 | 0.00 | 131.27 | $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** | **$23,848,830** | **$0.00791** |
| Spent Mercury Catalyst (lb) | 0 | 157 | 0.42 | $0 | $19,098 | $0.00001 |
| Flyash (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Slag (ton) | 0 | 587 | 16.23 | $0 | $2,781,394 | $0.00092 |
| *CO2 TS&M (1000 kg)* | *0* | *11,103* | *6.49* | *$0* | *$21,048,338* | *$0.00698* |
| | | | | | | |
| **By-products and Emissions** | | | | **$0** | **$0** | **$0.00000** |
| Sulfur (ton) | 0 | 50 | 0 | $0 | $0 | $0.00000 |
| | | | | | | |
| **TOTAL VARIABLE OPERATING COSTS** | | | | **$3,137,874** | **$54,142,033** | **$0.01795** |
| | | | | | | |
| **Fuel** | | | | **$0** | **$32,088,636** | **$0.01064** |
| Coal (ton) | 0 | 6,958 | 15.22 | $0 | $30,914,533 | $0.01025 |
| Natural Gas (MMBtu-HHV) | 0 | 804.18 | 5.00 | $0 | $1,174,103 | $0.00039 |

## Table T5-a9: Operating Cost for the Sour PSA with AGE and Claus plant case

**Operating and Maintenance Labor**

| **Operating Labor** | | |
|---|---|---|
| Operating Labor Rate | 34.65 | $/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 ($)** | **Annual Unit Cost ($/kW)** |
|---|---|---|
| **Fixed Operating Costs** | | |
| Operating Labor | $6,313,507 | $14,043.365 |
| Maintenance Labor | $13,682,329 | $30,434.104 |
| Administrative & Support Labor | $4,998,959 | $11,119.367 |
| Property Taxes & Insurance | $29,935,277 | $66,586.131 |
| | | |
| **TOTAL FIXED OPERATING COST** | **$54,930,072** | **$122,182.967** |

* (back calculated from NETL)

**Variable Operating Costs**

| | | | **Annual Cost ($)** | **Annual Unit Cost ($/kW.h)** |
|---|---|---|---|---|
| **Maintenance Material Cost** | 1.745% | of TPC* | **$ 26,118,098** | **$0.00829** |

| | **Consumption** | | | | | |
|---|---|---|---|---|---|---|
| | **Initial** | **per day** | **Unit Cost ($)** | **Initial Cost ($)** | **Annual Cost ($)** | **Annual Unit Cost ($/kW.h)** |
| **Water (1000 gallons)** | 0.00 | 2,938 | 1.08 | **$0** | **$928,025** | **$0.00029** |
| | | | | | | |
| **Chemicals** | | | | **$3,137,874** | **$5,051,099** | **$0.00160** |
| MU & WT Chem (lb) | 0 | 17,265 | 0.17 | $0 | $872,524 | $0.00028 |
| Carbon (Mercury Removal) (lb) | 114,477 | 157 | 1.05 | $120,220 | $48,088 | $0.00002 |
| COS Catalyst (m3) | 0 | 0 | 2397.36 | $0 | $0 | $0.00000 |
| Water Gas Shift Catalyst (ft3) | 6,049 | 4.14 | 498.83 | $3,017,654 | $603,531 | $0.00019 |
| AGE (Flexorb) Solution (Gal) | 0 | 0 | 0.00 | $0 | $3,500,000 | $0.00111 |
| 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 | $0 | $0 | $0.00000 |
| Claus Catalyst (ft3) | 0 | 0.70 | 131.27 | $0 | $26,956 | $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** | **$22,594,810** | **$0.00717** |
| Spent Mercury Catalyst (lb) | 0 | 157 | 0.42 | $0 | $19,098 | $0.00001 |
| Flyash (ton) | 0 | 0 | 0.00 | $0 | $0 | $0.00000 |
| Slag (ton) | 0 | 587 | 16.23 | $0 | $2,781,394 | $0.00088 |
| *CO2 TS&M (1000 kg)* | *0* | *10,442* | *6.49* | *$0* | *$19,794,318* | *$0.00628* |
| | | | | | | |
| **By-products and Emissions** | | | | **$0** | **$0** | **$0.00000** |
| Sulfur (ton) | 0 | 50 | 0 | $0 | $0 | $0.00000 |
| | | | | | | |
| **TOTAL VARIABLE OPERATING COSTS** | | | | **$3,137,874** | **$54,692,033** | **$0.01736** |
| | | | | | | |
| **Fuel** | | | | **$0** | **$30,914,533** | **$0.00981** |
| Coal (ton) | 0 | 6,958 | 15.22 | $0 | $30,914,533 | $0.00981 |
| Natural Gas (MMBtu-HHV) | 0 | 0.00 | 5.00 | $0 | $0 | $0.00000 |

## Table T5-a10: Cash flow summary for the Sour PSA with Oxy-combustor option.

| estimated COE | **129.37** | mills/kW.h |
|---|---|---|

| **end of** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | *1* | *2* | *3* | *4* | *5* | *6* | *7* | *8* | *9* | *10* | *11* | *12* | *13* | *14* | *15* | *16* | *17* | *18* |
| Operating revenues | $ - | $ - | $ - | $ - | $ - | $ 390,426 | $ 402,139 | $ 414,203 | $ 426,629 | $ 439,428 | $ 452,611 | $ 466,189 | $ 480,175 | $ 494,580 | $ 509,417 | $ 524,700 | $ 540,441 | $ 556,654 |
| Operating expenses | | | | | | | | | | | | | | | | | | |
| Fixed | $ - | $ - | $ - | $ - | $ - | $ 67,172 | $ 69,187 | $ 71,262 | $ 73,400 | $ 75,602 | $ 77,870 | $ 80,207 | $ 82,613 | $ 85,091 | $ 87,644 | $ 90,273 | $ 92,981 | $ 95,771 |
| Variable | $ - | $ - | $ - | $ - | $ - | $ 62,765 | $ 64,648 | $ 66,588 | $ 68,586 | $ 70,643 | $ 72,762 | $ 74,945 | $ 77,194 | $ 79,509 | $ 81,895 | $ 84,352 | $ 86,882 | $ 89,489 |
| Fuel | $ - | $ - | $ - | $ - | $ - | $ 37,200 | $ 38,316 | $ 39,465 | $ 40,649 | $ 41,868 | $ 43,124 | $ 44,418 | $ 45,751 | $ 47,123 | $ 48,537 | $ 49,993 | $ 51,493 | $ 53,038 |
| | | | | | | $ 167,137 | $ 172,151 | $ 177,315 | $ 182,635 | $ 188,114 | $ 193,757 | $ 199,570 | $ 205,557 | $ 211,724 | $ 218,075 | $ 224,618 | $ 231,356 | $ 238,297 |
| **Operating income** | $ - | $ - | $ - | $ - | $ - | $ 223,289 | $ 229,988 | $ 236,887 | $ 243,994 | $ 251,314 | $ 258,853 | $ 266,619 | $ 274,617 | $ 282,856 | $ 291,342 | $ 300,082 | $ 309,084 | $ 318,357 |
| Interest Expense | $ - | $ - | $ - | $ - | $ - | $ 48,076 | $ 45,931 | $ 43,667 | $ 41,280 | $ 38,760 | $ 36,102 | $ 33,298 | $ 30,340 | $ 27,219 | $ 23,927 | $ 20,453 | $ 16,788 | $ 12,922 |
| Depreciation & Amortization | $ - | $ - | $ - | $ - | $ - | $ 69,474.16 | $ 133,742.39 | $ 123,701.06 | $ 114,437.83 | $ 105,841.57 | $ 97,912.25 | $ 90,557.25 | $ 83,776.57 | $ 82,664.99 | $ 82,646.46 | $ 82,664.99 | $ 82,646.46 | $ 82,664.99 |
| **Taxable Income** | $ - | $ - | $ - | $ - | $ - | $ 105,739 | $ 50,315 | $ 69,519 | $ 88,277 | $ 106,712 | $ 124,839 | $ 142,763 | $ 160,501 | $ 172,972 | $ 184,768 | $ 196,964 | $ 209,650 | $ 222,770 |
| Income Taxes | $ - | $ - | $ - | $ - | $ - | $ 40,181 | $ 19,120 | $ 26,417 | $ 33,545 | $ 40,551 | $ 47,439 | $ 54,250 | $ 60,990 | $ 65,729 | $ 70,212 | $ 74,846 | $ 79,667 | $ 84,653 |
| **Net Income** | $ - | $ - | $ - | $ - | $ - | $ 65,558 | $ 31,195 | $ 43,102 | $ 54,732 | $ 66,161 | $ 77,400 | $ 88,513 | $ 99,510 | $ 107,242 | $ 114,556 | $ 122,118 | $ 129,983 | $ 138,117 |
| **Cash form Operation** | $ - | $ - | $ - | $ - | $ - | $ 223,289 | $ 229,988 | $ 236,887 | $ 243,994 | $ 251,314 | $ 258,853 | $ 266,619 | $ 274,617 | $ 282,856 | $ 291,342 | $ 300,082 | $ 309,084 | $ 318,357 |
| Income Taxes | $ - | $ - | $ - | $ - | $ - | $ 40,181 | $ 19,120 | $ 26,417 | $ 33,545 | $ 40,551 | $ 47,439 | $ 54,250 | $ 60,990 | $ 65,729 | $ 70,212 | $ 74,846 | $ 79,667 | $ 84,653 |
| Total Interest Expense | $ - | $ - | $ - | $ - | $ - | $ 48,076 | $ 45,931 | $ 43,667 | $ 41,280 | $ 38,760 | $ 36,102 | $ 33,298 | $ 30,340 | $ 27,219 | $ 23,927 | $ 20,453 | $ 16,788 | $ 12,922 |
| Total Principal Repayment | $ - | $ - | $ - | $ - | $ - | $ 39,008 | $ 41,153 | $ 43,417 | $ 45,805 | $ 48,324 | $ 50,982 | $ 53,786 | $ 56,744 | $ 59,865 | $ 63,157 | $ 66,631 | $ 70,296 | $ 74,162 |
| **Operating Cash Flow** | $ - | $ - | $ - | $ - | $ - | $ 96,024 | $ 123,784 | $ 123,386 | $ 123,365 | $ 123,679 | $ 124,330 | $ 125,285 | $ 126,543 | $ 130,043 | $ 134,045 | $ 138,151 | $ 142,333 | $ 146,620 |
| Capital Cost | $ 159,263 | $ 494,990 | $ 427,342 | $ 354,181 | $ 275,198 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow after Investment** | $ (159,263) | $ (494,990) | $ (427,342) | $ (354,181) | $ (275,198) | $ 96,024 | $ 123,784 | $ 123,386 | $ 123,365 | $ 123,679 | $ 124,330 | $ 125,285 | $ 126,543 | $ 130,043 | $ 134,045 | $ 138,151 | $ 142,333 | $ 146,620 |
| Loan Draws | $ 71,668.46 | $ 222,745.59 | $ 192,303.69 | $ 159,381.30 | $ 123,839.27 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow after Debt Financing** | $ (87,595) | $ (272,245) | $ (235,038) | $ (194,799) | $ (151,359) | $ 96,024 | $ 123,784 | $ 123,386 | $ 123,365 | $ 123,679 | $ 124,330 | $ 125,285 | $ 126,543 | $ 130,043 | $ 134,045 | $ 138,151 | $ 142,333 | $ 146,620 |
| Equity Draws | $ 87,595 | $ 272,245 | $ 235,038 | $ 194,799 | $ 151,359 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow for Equity Distribution** | $ - | $ - | $ - | $ - | $ - | $ 96,024 | $ 123,784 | $ 123,386 | $ 123,365 | $ 123,679 | $ 124,330 | $ 125,285 | $ 126,543 | $ 130,043 | $ 134,045 | $ 138,151 | $ 142,333 | $ 146,620 |

| **Internal rate of Return** | **12.00%** |
|---|---|
| Net Present Value at discount rate | |
| 8% | $ 528,039 |
| 10% | $ 209,876 |
| 12% | $ 0 |

estimated COE **128.70** mills/kW.h

| **end of** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** | **year** |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | *19* | *20* | *21* | *22* | *23* | *24* | *25* | *26* | *27* | *28* | *29* | *30* | *31* | *32* | *33* | *34* | *35* |
| Operating revenues | $ 573,353 | $ 590,554 | $ 608,271 | $ 626,519 | $ 645,314 | $ 664,674 | $ 684,614 | $ 705,152 | $ 726,307 | $ 748,096 | $ 770,539 | $ 793,655 | $ 817,465 | $ 841,989 | $ 867,249 | $ 893,266 | $ 920,064 |
| Operating expenses | | | | | | | | | | | | | | | | | |
| Fixed | $ 98,644 | $ 101,603 | $ 104,651 | $ 107,791 | $ 111,025 | $ 114,355 | $ 117,786 | $ 121,320 | $ 124,959 | $ 128,708 | $ 132,569 | $ 136,546 | $ 140,643 | $ 144,862 | $ 149,208 | $ 153,684 | $ 158,295 |
| Variable | $ 92,173 | $ 94,938 | $ 97,787 | $ 100,720 | $ 103,742 | $ 106,854 | $ 110,060 | $ 113,361 | $ 116,762 | $ 120,265 | $ 123,873 | $ 127,589 | $ 131,417 | $ 135,359 | $ 139,420 | $ 143,603 | $ 147,911 |
| Fuel | $ 54,629 | $ 56,268 | $ 57,956 | $ 59,694 | $ 61,485 | $ 63,330 | $ 65,230 | $ 67,186 | $ 69,202 | $ 71,278 | $ 73,416 | $ 75,619 | $ 77,888 | $ 80,224 | $ 82,631 | $ 85,110 | $ 87,663 |
| | $ 245,446 | $ 252,809 | $ 260,393 | $ 268,205 | $ 276,251 | $ 284,539 | $ 293,075 | $ 301,867 | $ 310,923 | $ 320,251 | $ 329,859 | $ 339,754 | $ 349,947 | $ 360,446 | $ 371,259 | $ 382,397 | $ 393,869 |
| **Operating income** | $ 327,908 | $ 337,745 | $ 347,877 | $ 358,314 | $ 369,063 | $ 380,135 | $ 391,539 | $ 403,285 | $ 415,384 | $ 427,845 | $ 440,680 | $ 453,901 | $ 467,518 | $ 481,543 | $ 495,990 | $ 510,869 | $ 526,196 |
| Interest Expense | $ 8,843 | $ 4,540 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| Depreciation & Amortization | $ 82,646.46 | $ 82,664.99 | $ 82,646.46 | $ 82,664.99 | $ 82,646.46 | $ 82,664.99 | $ 82,646.46 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Taxable Income** | $ 236,418 | $ 250,540 | $ 265,231 | $ 275,649 | $ 286,416 | $ 297,470 | $ 308,892 | $ 403,285 | $ 415,384 | $ 427,845 | $ 440,680 | $ 453,901 | $ 467,518 | $ 481,543 | $ 495,990 | $ 510,869 | $ 526,196 |
| Income Taxes | $ 89,839 | $ 95,205 | $ 100,788 | $ 104,746 | $ 108,838 | $ 113,039 | $ 117,379 | $ 153,248 | $ 157,846 | $ 162,581 | $ 167,459 | $ 172,482 | $ 177,657 | $ 182,987 | $ 188,476 | $ 194,130 | $ 199,954 |
| **Net Income** | $ 146,579 | $ 155,335 | $ 164,443 | $ 170,902 | $ 177,578 | $ 184,431 | $ 191,513 | $ 250,037 | $ 257,538 | $ 265,264 | $ 273,222 | $ 281,419 | $ 289,861 | $ 298,557 | $ 307,514 | $ 316,739 | $ 326,241 |
| **Cash form Operation** | $ 327,908 | $ 337,745 | $ 347,877 | $ 358,314 | $ 369,063 | $ 380,135 | $ 391,539 | $ 403,285 | $ 415,384 | $ 427,845 | $ 440,680 | $ 453,901 | $ 467,518 | $ 481,543 | $ 495,990 | $ 510,869 | $ 526,196 |
| Income Taxes | $ 89,839 | $ 95,205 | $ 100,788 | $ 104,746 | $ 108,838 | $ 113,039 | $ 117,379 | $ 153,248 | $ 157,846 | $ 162,581 | $ 167,459 | $ 172,482 | $ 177,657 | $ 182,987 | $ 188,476 | $ 194,130 | $ 199,954 |
| Total Interest Expense | $ 8,843 | $ 4,540 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| Total Principal Repayment | $ 78,241 | $ 82,544 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Operating Cash Flow** | $ 150,985 | $ 155,455 | $ 247,090 | $ 253,567 | $ 260,225 | $ 267,096 | $ 274,160 | $ 250,037 | $ 257,538 | $ 265,264 | $ 273,222 | $ 281,419 | $ 289,861 | $ 298,557 | $ 307,514 | $ 316,739 | $ 326,241 |
| Capital Cost | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow after Investment** | $ 150,985 | $ 155,455 | $ 247,090 | $ 253,567 | $ 260,225 | $ 267,096 | $ 274,160 | $ 250,037 | $ 257,538 | $ 265,264 | $ 273,222 | $ 281,419 | $ 289,861 | $ 298,557 | $ 307,514 | $ 316,739 | $ 326,241 |
| Loan Draws | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow after Debt Financing** | $ 150,985 | $ 155,455 | $ 247,090 | $ 253,567 | $ 260,225 | $ 267,096 | $ 274,160 | $ 250,037 | $ 257,538 | $ 265,264 | $ 273,222 | $ 281,419 | $ 289,861 | $ 298,557 | $ 307,514 | $ 316,739 | $ 326,241 |
| Equity Draws | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow for Equity Distribution** | $ 150,985 | $ 155,455 | $ 247,090 | $ 253,567 | $ 260,225 | $ 267,096 | $ 274,160 | $ 250,037 | $ 257,538 | $ 265,264 | $ 273,222 | $ 281,419 | $ 289,861 | $ 298,557 | $ 307,514 | $ 316,739 | $ 326,241 |

| | |
|---|---|
| **Internal rate of Return** | **12.00%** |
| Net Present Value at discount rate | |
| 8% | $ 528,039 |
| 10% | $ 209,876 |
| 12% | $ 0 |

## Table T5-a11: Cash flow summary for the Sour PSA with AGE and Claus plant option

| estimated COE | 118.26 | mills/kW.h |
|---|---|---|

| end of | year | year | year | year | year | year | year | year | year | year | year | year | year | year | year | year | year | year |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 |
| Operating revenues | $ - | $ - | $ - | $ - | $ - | $ 372,848 | $ 384,033 | $ 395,554 | $ 407,421 | $ 419,644 | $ 432,233 | $ 445,200 | $ 458,556 | $ 472,313 | $ 486,482 | $ 501,077 | $ 516,109 | $ 531,592 |
| Operating expenses | | | | | | | | | | | | | | | | | | |
| Fixed | $ - | $ - | $ - | $ - | $ - | $ 63,679 | $ 65,589 | $ 67,557 | $ 69,584 | $ 71,671 | $ 73,821 | $ 76,036 | $ 78,317 | $ 80,667 | $ 83,087 | $ 85,579 | $ 88,147 | $ 90,791 |
| Variable | $ - | $ - | $ - | $ - | $ - | $ 63,403 | $ 65,305 | $ 67,264 | $ 69,282 | $ 71,361 | $ 73,502 | $ 75,707 | $ 77,978 | $ 80,317 | $ 82,727 | $ 85,208 | $ 87,765 | $ 90,398 |
| Fuel | $ - | $ - | $ - | $ - | $ - | $ 35,838 | $ 36,914 | $ 38,021 | $ 39,162 | $ 40,336 | $ 41,547 | $ 42,793 | $ 44,077 | $ 45,399 | $ 46,761 | $ 48,164 | $ 49,609 | $ 51,097 |
| | | | | | | $ 162,920 | $ 167,808 | $ 172,842 | $ 178,028 | $ 183,368 | $ 188,869 | $ 194,536 | $ 200,372 | $ 206,383 | $ 212,574 | $ 218,952 | $ 225,520 | $ 232,286 |
| **Operating income** | $ - | $ - | $ - | $ - | $ - | $ 209,927 | $ 216,225 | $ 222,712 | $ 229,393 | $ 236,275 | $ 243,364 | $ 250,664 | $ 258,184 | $ 265,930 | $ 273,908 | $ 282,125 | $ 290,589 | $ 299,306 |
| Interest Expense | $ - | $ - | $ - | $ - | $ - | $ 45,232 | $ 43,214 | $ 41,084 | $ 38,838 | $ 36,467 | $ 33,967 | $ 31,329 | $ 28,545 | $ 25,609 | $ 22,511 | $ 19,243 | $ 15,795 | $ 12,158 |
| Depreciation & Amortization | $ - | $ - | $ - | $ - | $ - | $ 65,297.69 | $ 125,702.40 | $ 116,264.71 | $ 107,558.35 | $ 99,478.85 | $ 92,026.21 | $ 85,113.36 | $ 78,740.30 | $ 77,695.54 | $ 77,678.13 | $ 77,695.54 | $ 77,678.13 | $ 77,695.54 |
| **Taxable Income** | $ - | $ - | $ - | $ - | $ - | $ 99,397 | $ 47,309 | $ 65,363 | $ 82,997 | $ 100,329 | $ 117,370 | $ 134,222 | $ 150,899 | $ 162,625 | $ 173,718 | $ 185,186 | $ 197,115 | $ 209,453 |
| Income Taxes | $ - | $ - | $ - | $ - | $ - | $ 37,771 | $ 17,977 | $ 24,838 | $ 31,539 | $ 38,125 | $ 44,601 | $ 51,004 | $ 57,341 | $ 61,798 | $ 66,013 | $ 70,371 | $ 74,904 | $ 79,592 |
| **Net Income** | $ - | $ - | $ - | $ - | $ - | $ 61,626 | $ 29,332 | $ 40,525 | $ 51,458 | $ 62,204 | $ 72,770 | $ 83,218 | $ 93,557 | $ 100,828 | $ 107,705 | $ 114,815 | $ 122,212 | $ 129,861 |
| **Cash form Operation** | $ - | $ - | $ - | $ - | $ - | $ 209,927 | $ 216,225 | $ 222,712 | $ 229,393 | $ 236,275 | $ 243,364 | $ 250,664 | $ 258,184 | $ 265,930 | $ 273,908 | $ 282,125 | $ 290,589 | $ 299,306 |
| Income Taxes | $ - | $ - | $ - | $ - | $ - | $ 37,771 | $ 17,977 | $ 24,838 | $ 31,539 | $ 38,125 | $ 44,601 | $ 51,004 | $ 57,341 | $ 61,798 | $ 66,013 | $ 70,371 | $ 74,904 | $ 79,592 |
| Total Interest Expense | $ - | $ - | $ - | $ - | $ - | $ 45,232 | $ 43,214 | $ 41,084 | $ 38,838 | $ 36,467 | $ 33,967 | $ 31,329 | $ 28,545 | $ 25,609 | $ 22,511 | $ 19,243 | $ 15,795 | $ 12,158 |
| Total Principal Repayment | $ - | $ - | $ - | $ - | $ - | $ 36,700 | $ 38,719 | $ 40,848 | $ 43,095 | $ 45,465 | $ 47,966 | $ 50,604 | $ 53,387 | $ 56,324 | $ 59,421 | $ 62,690 | $ 66,138 | $ 69,775 |
| **Operating Cash Flow** | $ - | $ - | $ - | $ - | $ - | $ 90,224 | $ 116,315 | $ 115,941 | $ 115,922 | $ 116,217 | $ 116,830 | $ 117,727 | $ 118,910 | $ 122,199 | $ 125,962 | $ 129,821 | $ 133,752 | $ 137,781 |
| Capital Cost | $ 149,676 | $ 465,194 | $ 401,618 | $ 332,861 | $ 258,633 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow after Investment** | $ (149,676) | $ (465,194) | $ (401,618) | $ (332,861) | $ (258,633) | $ 90,224 | $ 116,315 | $ 115,941 | $ 115,922 | $ 116,217 | $ 116,830 | $ 117,727 | $ 118,910 | $ 122,199 | $ 125,962 | $ 129,821 | $ 133,752 | $ 137,781 |
| Loan Draws | $ 67,354.37 | $ 209,337.39 | $ 180,727.95 | $ 149,787.32 | $ 116,384.75 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow after Debt Financing** | $ (82,322) | $ (255,857) | $ (220,890) | $ (183,073) | $ (142,248) | $ 90,224 | $ 116,315 | $ 115,941 | $ 115,922 | $ 116,217 | $ 116,830 | $ 117,727 | $ 118,910 | $ 122,199 | $ 125,962 | $ 129,821 | $ 133,752 | $ 137,781 |
| Equity Draws | $ 82,322 | $ 255,857 | $ 220,890 | $ 183,073 | $ 142,248 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow for Equity Distribution** | $ - | $ - | $ - | $ - | $ - | $ 90,224 | $ 116,315 | $ 115,941 | $ 115,922 | $ 116,217 | $ 116,830 | $ 117,727 | $ 118,910 | $ 122,199 | $ 125,962 | $ 129,821 | $ 133,752 | $ 137,781 |

| | |
|---|---|
| **Internal rate of Return** | **12.00%** |
| Net Present Value at discount rate | |
| 8% | $ 496,335 |
| 10% | $ 197,271 |
| 12% | $ (0) |

| estimated COE | 117.82 | mills/kW.h |
|---|---|---|

| end of | year | year | year | year | year | year | year | year | year | year | year | year | year | year | year | year | year |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | 19 | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 | 34 | 35 |
| Operating revenues | $ 547,540 | $ 563,966 | $ 580,885 | $ 598,312 | $ 616,261 | $ 634,749 | $ 653,791 | $ 673,405 | $ 693,607 | $ 714,415 | $ 735,848 | $ 757,923 | $ 780,661 | $ 804,081 | $ 828,203 | $ 853,049 | $ 878,641 |
| Operating expenses | | | | | | | | | | | | | | | | | |
| Fixed | $ 93,515 | $ 96,320 | $ 99,210 | $ 102,186 | $ 105,252 | $ 108,409 | $ 111,662 | $ 115,011 | $ 118,462 | $ 122,016 | $ 125,676 | $ 129,446 | $ 133,330 | $ 137,330 | $ 141,449 | $ 145,693 | $ 150,064 |
| Variable | $ 93,110 | $ 95,903 | $ 98,780 | $ 101,743 | $ 104,796 | $ 107,939 | $ 111,178 | $ 114,513 | $ 117,948 | $ 121,487 | $ 125,131 | $ 128,885 | $ 132,752 | $ 136,734 | $ 140,837 | $ 145,062 | $ 149,413 |
| Fuel | $ 52,630 | $ 54,209 | $ 55,835 | $ 57,510 | $ 59,235 | $ 61,013 | $ 62,843 | $ 64,728 | $ 66,670 | $ 68,670 | $ 70,730 | $ 72,852 | $ 75,038 | $ 77,289 | $ 79,607 | $ 81,996 | $ 84,456 |
| | $ 239,254 | $ 246,432 | $ 253,825 | $ 261,440 | $ 269,283 | $ 277,361 | $ 285,682 | $ 294,253 | $ 303,080 | $ 312,172 | $ 321,538 | $ 331,184 | $ 341,119 | $ 351,353 | $ 361,893 | $ 372,750 | $ 383,933 |
| **Operating income** | $ 308,286 | $ 317,534 | $ 327,060 | $ 336,872 | $ 346,978 | $ 357,388 | $ 368,109 | $ 379,152 | $ 390,527 | $ 402,243 | $ 414,310 | $ 426,739 | $ 439,542 | $ 452,728 | $ 466,310 | $ 480,299 | $ 494,708 |
| Interest Expense | $ 8,320 | $ 4,271 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| Depreciation & Amortization | $ 77,678.13 | $ 77,695.54 | $ 77,678.13 | $ 77,695.54 | $ 77,678.13 | $ 77,695.54 | $ 77,678.13 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Taxable Income** | $ 222,287 | $ 235,567 | $ 249,382 | $ 259,176 | $ 269,300 | $ 279,692 | $ 290,431 | $ 379,152 | $ 390,527 | $ 402,243 | $ 414,310 | $ 426,739 | $ 439,542 | $ 452,728 | $ 466,310 | $ 480,299 | $ 494,708 |
| Income Taxes | $ 84,469 | $ 89,516 | $ 94,765 | $ 98,487 | $ 102,334 | $ 106,283 | $ 110,364 | $ 144,078 | $ 148,400 | $ 152,852 | $ 157,438 | $ 162,161 | $ 167,026 | $ 172,037 | $ 177,198 | $ 182,514 | $ 187,989 |
| **Net Income** | $ 137,818 | $ 146,052 | $ 154,617 | $ 160,689 | $ 166,966 | $ 173,409 | $ 180,067 | $ 235,074 | $ 242,127 | $ 249,391 | $ 256,872 | $ 264,578 | $ 272,516 | $ 280,691 | $ 289,112 | $ 297,785 | $ 306,719 |
| **Cash form Operation** | $ 308,286 | $ 317,534 | $ 327,060 | $ 336,872 | $ 346,978 | $ 357,388 | $ 368,109 | $ 379,152 | $ 390,527 | $ 402,243 | $ 414,310 | $ 426,739 | $ 439,542 | $ 452,728 | $ 466,310 | $ 480,299 | $ 494,708 |
| Income Taxes | $ 84,469 | $ 89,516 | $ 94,765 | $ 98,487 | $ 102,334 | $ 106,283 | $ 110,364 | $ 144,078 | $ 148,400 | $ 152,852 | $ 157,438 | $ 162,161 | $ 167,026 | $ 172,037 | $ 177,198 | $ 182,514 | $ 187,989 |
| Total Interest Expense | $ 8,320 | $ 4,271 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| Total Principal Repayment | $ 73,613 | $ 77,661 | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Operating Cash Flow** | $ 141,884 | $ 146,086 | $ 232,295 | $ 238,385 | $ 244,644 | $ 251,105 | $ 257,745 | $ 235,074 | $ 242,127 | $ 249,391 | $ 256,872 | $ 264,578 | $ 272,516 | $ 280,691 | $ 289,112 | $ 297,785 | $ 306,719 |
| Capital Cost | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow after Investment** | $ 141,884 | $ 146,086 | $ 232,295 | $ 238,385 | $ 244,644 | $ 251,105 | $ 257,745 | $ 235,074 | $ 242,127 | $ 249,391 | $ 256,872 | $ 264,578 | $ 272,516 | $ 280,691 | $ 289,112 | $ 297,785 | $ 306,719 |
| Loan Draws | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow after Debt Financing** | $ 141,884 | $ 146,086 | $ 232,295 | $ 238,385 | $ 244,644 | $ 251,105 | $ 257,745 | $ 235,074 | $ 242,127 | $ 249,391 | $ 256,872 | $ 264,578 | $ 272,516 | $ 280,691 | $ 289,112 | $ 297,785 | $ 306,719 |
| Equity Draws | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - | $ - |
| **Net Cash Flow for Equity Distribution** | $ 141,884 | $ 146,086 | $ 232,295 | $ 238,385 | $ 244,644 | $ 251,105 | $ 257,745 | $ 235,074 | $ 242,127 | $ 249,391 | $ 256,872 | $ 264,578 | $ 272,516 | $ 280,691 | $ 289,112 | $ 297,785 | $ 306,719 |

| **Internal rate of Return** | **12.00%** |
|---|---|
| Net Present Value at discount rate | |
| 8% | $ 496,335 |
| 10% | $ 197,271 |
| 12% | $ (0) |