Final Report

Project Title: A Low Cost, High Capacity Regenerable Sorbent for Pre-combustion CO_2 Capture

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A Low Cost, High Capacity Regenerable Sorbent for Pre-combustion CO_2 Capture

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1. Executive Summary

The objective of this project is to develop a new pre-combustion carbon capture technology and demonstrate its technical feasibility and economic viability in laboratory-scale tests, in field demonstrations, and by carrying out a detailed process design and analysis of the new system as part of an Integrated Gasification Combined Cycle (IGCC) power plant. The new technology uses a low cost, high capacity adsorbent that selectively removes CO_2 above the dew point of the synthesis gas (temperatures from 190 to 260°C are explored in this study depending on the gasifier type).

The sorbent is based on a TDA proprietary mesoporous carbon that consists of surface functionalized groups that remove CO_2 via physical adsorption. The relatively strong interaction between the surface group sand CO_2 enables effective operation at high temperatures. However, because the sorbent and the CO_2 do not form a true covalent bond, the energy needed to regenerate the sorbent is much lower (less than 5 kcal per mol of CO_2) than that observed for either chemical absorbents (e.g., 29.9 kcal/mol CO_2 for sodium carbonate) or amine-based solvents (e.g., 14.2 kcal/mol CO_2 for mono-ethanolamine). The sorbent is regenerated using a combination of pressure swing (i.e., recovering CO_2 as a concentrated gas at pressure to 9.7 barA) and concentration swing (using 2.4 bed volumes of steam purge per regeneration) while maintaining near isothermal operation in the sorbent beds. Avoiding temperature swing eliminates the long heating/cooling transitions and enables the rapid adsorption and regeneration that leads to short cycle times, increasing utilization and reducing the overall sorbent inventory. The high surface area and favorable porosity of the sorbent also provide a unique platform to introduce additional functionality, such as active groups to catalyze the water-gas-shift (WGS) reaction or to remove trace metals (e.g., Hg, As).

In the project, we first optimized the sorbent formulation and developed a process to manufacture the sorbent using high throughput production equipment. The optimized sorbent uses widely available, low cost raw materials and can be mass-produced in large volumes. TDA optimized the sorbent formulation to increase its mechanical integrity (i.e., crush strength) to ensure long on-stream life and minimum dusting during loading and handling. The CO_2 adsorption isotherms generated from bench-scale fixed bed breakthrough experiments showed that the sorbent can achieve over 20% wt. capacity at an adsorption temperature of 180°C and a CO_2 partial pressure of 600 psia (7.2%wt. at 198°C and 200 psia CO_2 partial pressure at the operating conditions of a typical state-of-the-art oxygen-blown coal gasifier). These results suggest a very high working capacity for the adsorbent, even if the CO_2 is recovered (i.e., sorbent regeneration is carried out) at high pressure.

In a laboratory-scale reactor, we have demonstrated the cyclic activity and life of the material for over 11,650 adsorption/regeneration cycles under representative conditions. In these cycles, the sorbent maintained its CO_2 adsorption capacity and removal efficiency without any signs of degradation (Figure 1). Extensive characterization of the sorbent following the multiple cycle tests suggests that the physical and chemical characteristics of the sorbent (e.g., surface area, active material content) remained intact.

Based on the bench-scale test results, TDA designed a pressure swing adsorption (PSA) cycle sequence that consists of 8 PSA steps, including adsorption, depressurization, pressure equalization, desorption and steam purge all of which ensures continuous removal of CO_2 from the synthesis gas. The system generates a concentrated CO_2 stream ready for further purification and compression and also a very low CO_2 containing synthesis gas fuel to be burned in the gas turbine. The PSA cycle sequence was optimized using a Fortran code developed by University of Michigan. Key operating trade-offs such as the CO_2 recovery pressure and steam consumption in the purge step have been optimized and all critical operating parameters have been determined. Based on the simulation results, TDA completed the design of the CO_2 separation, purification and compression process. For a nominal 600 MW IGCC plant, the CO_2 removal system consists of three trains of 8-bed PSA units; total of 24 beds housing the sorbent.

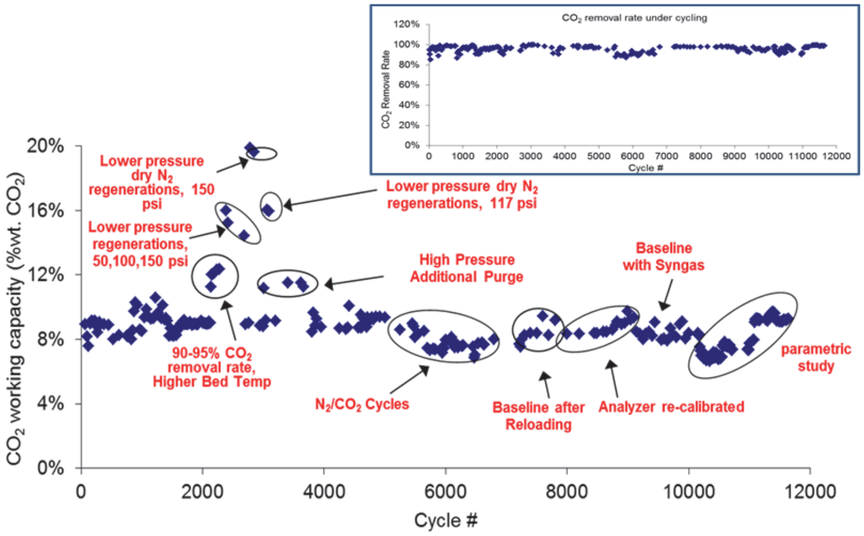


Figure 1. The results of the multiple-cycle experiments. T= 240°C, P= 500 psia, CO_2 = 40% vol., simulated synthesis gas.

In collaboration with the Southern Company and Phillips 66, TDA carried out two slipstream demonstrations to assess the efficacy of the sorbent for removing CO_2 from actual coal-derived synthesis gas. We first built a fully automated test unit with 4 high temperature PSA beds to house the sorbent (although the slipstream test unit did not use the 8-bed configuration selected for the final design, the selected configuration was more than adequate for demonstrating all critical aspects of the concept within the budget and schedule constraints of this project). The field demonstration unit also included an additional synthesis gas conditioning sub-assembly, with options for bulk sulfur removal, steam injection and water-gas-shift synthesis to adjust the synthesis gas concentration and purity. Figure 2 shows the pictures of the field demonstration unit, including the CO_2 separation and synthesis gas conditioning sub-assemblies.

Figure 2. TDA's field demonstration unit; (a) the 4-bed high temperature PSA-based CO_2 separation sub-assembly; (b) synthesis gas conditioning sub-assembly, (c) TDA's field demonstration unit installed at the Power Systems Demonstration Facility, (d) TDA's field demonstration unit installed at the Wabash River IGCC power Plant.

The field testing results at the Power Systems Demonstration Facility, Wilsonville, Alabama showed the pre-combustion capture technology is fully capable of removing CO_2 from the synthesis gas generated by an air-blown transport gasifier. The sorbent performance in the field closely matched the results in the laboratory using simulated synthesis gas mixture (with high nitrogen content and low overall pressure) suggesting that the potential impurities in the coal-derived synthesis gas did not lead to degradation in sorbent performance. The sorbent maintained its performance for over 1,030 adsorption/regeneration cycles.

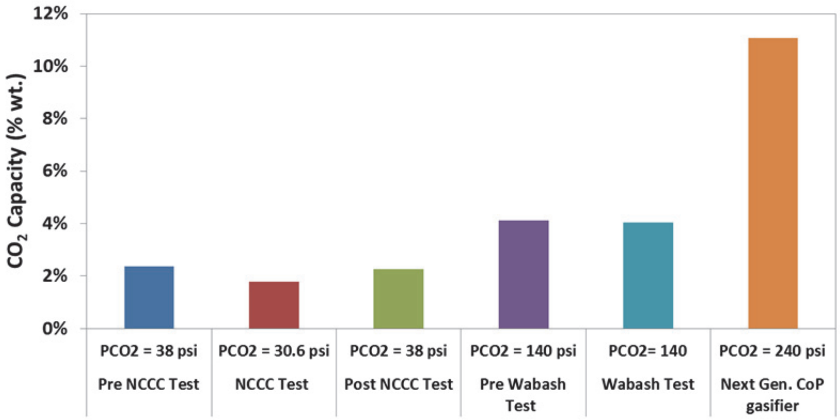


Figure 3. Sorbent performance comparison measured in laboratory tests using simulated synthesis gas and in the field tests with actual coal-derived synthesis gas.

The evaluation of the unit in Wabash River IGCC Power Plant, Terre Haute, Indiana was also successful. The sorbent removed CO_2 with much higher capacity due to the high CO_2 partial pressure provided by the oxygen-blown E-Gas^{TM} gasifier operating at higher pressure. Sorbent performance was demonstrated for an additional 715 cycles under a range of conditions (the same sorbent was used in PSDF and Wabash River demonstrations).

Figure 3 shows the average CO_2 capacity of the sorbent measured in laboratory tests using simulated synthesis gas and in the field tests with actual coal-derived synthesis gas. The similar CO_2 capacity observed before, during and after testing of the sorbent at the PSDF suggest minimal impact of synthesis gas impurities in the operation of the sorbent (altogether 26,750 SCF of synthesis gas has been treated). It is speculated that the high temperature operation of the CO_2 separation system prevented the fouling of the sorbent by the tars (e.g., benzene, toluene, xylene) and alkali compounds (NaCl, KCl vapors). The CO_2 capacity demonstrated at the Wabash River demonstration and in lab-scale tests at TDA were also in good agreement. A slight decrease in capacity in the Wabash River demonstration is attributed to the non-optimal operation of the WGS catalyst (which converts the CO into CO_2), and the related lower CO_2 partial pressure that will drive the adsorption process.

In collaboration with the Advanced Power and Energy Program of the University of California, Irvine (UCI), TDA developed system simulation models using Aspen Plus^{TM} simulation software to assess the economic viability of TDA's high temperature PSA-based pre-combustion carbon capture technology. The analysis provided a thorough comparison against a conventional cold gas cleanup technology based on the Selexol™ physical solvent scrubbing. In the simulation work, UCI strictly followed the guidelines and assumptions established by DOE/NETL for assessing the economic viability of competing CO_2 capture technologies. First, the accuracy of the Aspen Model was verified by checking the simulation results against a previous DOE/NETL analysis (Case 4 in the updated DOE/NETL study report DOE/NETL-2010/1397). The results of the UCI's Aspen model simulating the cold gas cleanup with Selexol™ were in close agreement with the results of the DOE/NETL.

The analysis explored several design options in which the high temperature PSA system is integrated with various commercial gasifiers, including Phillips 66's E-Gas^{TM} gasification technology. The simulation results showed that the CO_2 product purity from TDA's high temperature PSA system at 90% carbon capture could match the purity levels that can be achieved by the Selexol^{TM} process (with an option to further reducing the concentration of Ar and N_2 to less than 20 ppmv, if desired). The TDA's pre-combustion CO_2 capture system achieves a net plant efficiency of 34% on a coal high heating value (HHV) basis. This net plant efficiency is significantly higher than that can be achieved for the Selexol^{TM} scrubber at 31.6%, corresponding to a 7.5% decrease in the heat rate for TDA's Warm Gas Cleanup system. The water consumption in the plant on a net kW generated basis is also significantly lower for the TDA's Warm gas Cleanup process, a reduction of 8.5 kmol per MWh, thereby preserving a valuable resource.

The capital expense for the plant was estimated following the cost guidelines provided by DOE/NETL (2007) (e.g. higher contingencies were applied for unproven technologies such as the high temperature PSA unit.) The total plant cost for TDA process is estimated as $2,418/kW, which is 12% lower than that of the Cold Gas Cleanup Case at $2,754/kW. The levelized cost of electricity including the transport, storage and monitoring (TS&M) costs for CO_2 is calculated as $92.9/MWh for TDA's Warm gas Cleanup Case. This cost for the Selexol^{TM} scrubbing technology is calculated as $105.2/MWh. Table 1 summarizes the system analysis results.

The results of the system analysis suggest that TDA's high temperature PSA-based Warm Gas Clean-up Technology can make a substantial improvement in the IGCC plant thermal performance for achieving near zero emissions, including greater than 90% carbon capture. The capital expenses are also expected to be lower than that of Selexol's™. The higher net plant efficiency and lower capital and operating costs results in substantial reduction in the cost of electricity for the IGCC plant equipped with TDA's high temperature PSA-based carbon capture system.

Table 1. Comparison of TDA's Warm Gas Cleanup against SelexolTM based Cold Gas Cleanup System. Basis: IGCC plant operating with fully loaded GE F class gas turbine that generates 464 MW_e power.

	Cold Gas Cleanup Selexol^{TM}	**Warm Gas Cleanup TDA's CO_2 Sorbent**
CO_2 Capture, %	90.0	90.0
Gross Power Generated, kWe	691,247	733,028
Gas Turbine Power	464,000	464,000
Steam Turbine Power	227,247	269,028
Auxiliary Load, kWe	175,994	131,163
Net Power, kWe	515,253	601,865
Net Plant Efficiency, % HHV	**31.6%**	**34.0%**
Coal Feed Rate, kg/h	216,187	234,867
Raw Water Usage, GPM/MWe	11.8	11.2
Total Plant Cost, $/kWe	2,754	2,418
COE without CO_2 TS&M, $/MWh	99.8	87.8
COE with CO2 TS&M, $/MWh	**105.2**	**92.9**

The results of the DE-FE0000469 project suggest that the high temperature PSA-based pre-combustion carbon capture technology merits further research and development. It is recommended that the performance of a fully-equipped system should be demonstrated at larger-scale (e.g., 1 MW) using actual synthesis gas for a longer duration (6 to 12 months). The system should contain all critical components (including all PSA reactors, accumulators etc.) to fully demonstrate the cycle sequence, enabling the demonstration of product purity (both the hydrogen-rich fuel gas and CO_2-rich retentate). The demonstration duration should be long enough to allow over 20,000 cycles (with a 16 min full cycle time it corresponds to 7.5 months of testing) under optimum operating conditions. The selected demonstration scale should be large enough require sorbent production in large quantities (at the recommended 1 MW demonstration 8 m^3 of sorbent will be needed). A more detailed system simulation and cost analysis is also recommended, including design work and accurate quotes from the suppliers of the major process equipment (e.g., air separation unit, gasifier, CO_2 compressors). Successful completion of this recommended work will provide the basis for the new technology to be employed in potential commercial pilot-scale demonstrations (50-100 MW scale).

2. Introduction

Coal accounts for 56% of U.S. power generation and its contribution to future energy supply is expected to increase since U.S. has 25% of world's coal reserves (Tonks 2007). Coal-fueled Integrated Gasification Combined Cycle (IGCC) systems are environmentally superior to pulverized coal (PC)-fired boilers not only because they are more efficient at producing electricity, but also they can be equipped with more cost effective technologies for CO_2 capture and pollution control.

TDA has developed a low cost, high capacity CO_2 sorbent and demonstrated its technical and economic viability for pre-combustion CO_2 capture. The sorbent material removes CO_2 via strong physical adsorption and this relatively strong interaction enables effective operation at temperatures up to 300°C (well above the synthesis gas dew point). However, because the sorbent and the CO_2 do not form a true covalent bond, the energy needed to regenerate our sorbent (~5.0 kcal per mol of CO_2) is much lower than that observed for either chemical absorbents (e.g., 29.9 kcal/mol CO_2 for sodium carbonate) or amine-based solvents (e.g., 14.2 kcal/mol CO_2 for monoethanolamine). TDA's sorbent is used in a high temperature pressure swing adsorption (PSA) system to selectively remove CO_2 above the dew point of the synthesis gas (180 to 260°C is explored in this study based on the gasifier type). The sorbent is regenerated using a combination of pressure swing (i.e., reducing the pressure to 9.7 barA) and concentration swing (using 2.4 bed volumes of steam purge per regeneration) while maintaining near isothermal operation. The system undergo a PSA sequence, including adsorption, depressurization, pressure equalization, desorption and steam purge all of which ensures continuous removal of CO_2 from the synthesis gas, generating a concentrated CO_2 stream ready for further purification and compression and also a very low CO_2 containing synthesis gas fuel to be burned in the gas turbine.

2.1 IGCC Plant Integrated with TDA's Sorbent Based CO_2 Capture Process

Figure 4 shows a simplified block diagram of an IGCC plant integrated with TDA's high temperature PSA-based CO_2 capture process. The new pre-combustion carbon capture system can be integrated with any type of gasifier (air-blown versus oxygen-blow or transport gasifier versus water-quenched gasifier). In each case the performance of the CO_2 separation and purification system will depend on the CO_2 partial pressure in the raw synthesis gas.

In our process, following the gasification and particulate control units, the bulk of the sulfur in the synthesis gas is removed with a warm gas desulfurization technology (such as the regenerable zinc-titanate-based syngas desulfurization system developed by the Research Triangle Institute or the Z-Sorb technology commercialized by Phillips 66). Warm gas desulfurization will be needed for any high temperature pre-combustion control technology as the objective is not to cool the synthesis gas down to low temperatures for sulfur control. TDA's Warm gas CO_2 capture system is located downstream of a warm gas desulfurization system (we selected RTI's system for our process design). The regeneration off-gas from the desulfurization system is further treated in a sulfuric acid plant, ultimately converting all sulfur into a concentrated H_2SO_4 product. The desulfurized synthesis gas is then fed to the water-gas-shift system that converts the CO into H_2; multiple stages of water-gas-shift reactors with inter-bed cooling was selected to ensure the highest CO conversion into CO_2 (as in other carbon capture processes, since the sorbent is selective for CO_2, high conversion of CO to CO_2 is essential to achieve high carbon capture).

Consistent with previous DOE analysis, we maintained a H_2O:CO molar ratio of 2.0 at the inlet of the high temperature shift reactor (1 more mole of H_2O than is required by reaction stoichiometry). The synthesis gas from the WGS unit is sent to the CO_2 Capture system at a temperature slightly (at least 30°C) above the dew point of the synthesis gas. This eliminates the need to cool the synthesis gas below its dew point using condensing heat exchangers.

The CO_2 capture block consists of a CO_2 separation system (the high temperature PSA system) and a purification/compression system, which further treats the CO_2 stream from the separation unit into a pure, pressurized CO_2 product that meets pipeline specification. As required by the FOA, TDA's high temperature PSA-based CO_2 separation system captures 90% of the carbon from the synthesis gas as CO_2 and produces a CO_2-lean synthesis gas (primarily H_2 and H_2O) that is sent to the gas turbine. Any gases trapped in the voids of the sorbent and the reactor ullage space are recovered at an intermediate pressure and recycled back to the synthesis gas feed to ensure high H_2 recovery in the CO_2 separation unit. A steam purge at lower pressure is used to fully regenerate the sorbent.

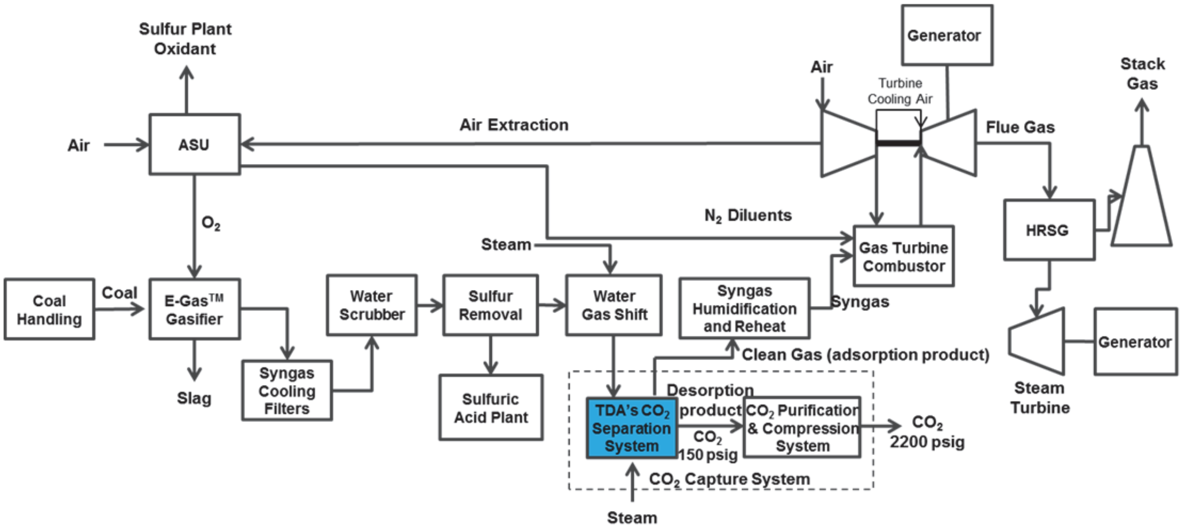


Figure 4. TDA's CO_2 separation system Integrated with an IGCC power plant.

A CO_2 rich stream primarily consisting of H_2O and CO_2 (along with some CO and H_2 impurities) is sent for further purification and compression to produce high purity CO_2 at 2,200 psig that can be sent for sequestration. For CO_2 purification, two approaches have been explored:

1. Cryogenic separation of CO_2 – In this approach after the removal of water the CO_2-rich off-gases are cooled to low temperatures to recover CO_2 as a pure liquid (while generating a overhead stream for all the other gases trapped within the sorbent pores, along with CO_2 to be recycled back to the process)
2. Catalytic combustor to burn any residual syngas with oxygen – In this approach pure oxygen is used to oxidize any H_2, CO and CH_4 remaining in the concentrated CO_2 stream, converting them into CO_2 and H_2O. After dehydration, CO_2 was pressurized.

As it will be discussed in the System Analysis, both approaches proved more economical than the Selexol^{TM}-based state-of-the-art pre-combustion carbon capture technology. The first option provided a very pure CO_2 stream (with Ar and N_2 content in CO_2 less than 20 ppmv). The catalytic combustion of impurities provided CO_2 product purity comparable to that can be achieved by Selexol^{TM}, while it resulted in lower system complexity and lower capital and operating costs.

3. Project Objectives

The overall objective of the proposed research is to develop a low cost, high capacity CO_2 sorbent and demonstrate its technical and economic viability for pre-combustion CO_2 capture. The specific objectives supporting our research plan were to optimize the chemical structure and physical properties of the sorbent, scale-up its production using high throughput manufacturing equipment and bulk raw materials and then evaluate its performance, first in bench-scale experiments and then in slipstream tests using actual coal-derived synthesis gas. One of the objectives of the laboratory-scale evaluations was to demonstrate the life and durability of the sorbent for over 10,000 cycles and to assess the impact of contaminants (such as sulfur) on its performance. In the field tests, our objective was to demonstrate the operation of the sorbent using actual coal-derived synthesis gas streams generated by air-blown and oxygen-blown commercial and pilot-scale coal gasifiers (the CO_2 partial pressure in these gas streams is significantly different, which directly impacts the operating conditions hence the performance of the sorbent).

To support the field demonstration work, TDA collaborated with Phillips 66 and Southern Company to carry out two separate field tests using actual coal-derived synthesis gas at the Wabash River IGCC Power Plant in Terre Haute, IN and the National Carbon Capture Center (NCCC) in Wilsonville, AL. In collaboration with the University of California, Irvine (UCI), a detailed engineering and economic analysis for the new CO_2 capture system was also proposed to be carried out using Aspen Plus^{TM} simulation software, and estimate its effect on the plant efficiency. All analyses were consistent with DOE's Cost Estimation Guidelines provided in the DOE/NETL Report (DOE/NETL-2010/1397) to compare all competing technologies (including the state-of-the-art Selexol^{TM} based solvent scrubbing technology) on a fair basis.

3.1 Work Plan

The R&D effort was divided into eleven tasks (the following description is taken from the proposal). In Task 1, we optimized the sorbent composition and production methodology. In Task 2, we evaluated the sorbent at bench-scale under representative operating conditions. In Task 3 TDA in collaboration with UCI carried out the process design. TDA provided a preliminary system design to UCI and UCI carried out the detailed process design and modeling using Aspen Plus^{TM} for material and energy balances and calculated the net plant efficiency. In Task 4, we scaled-up the sorbent production using scalable production equipment and carried out a detailed cost analysis. In Task 5, we carried out long–term cycling experiments and evaluated the impact of contaminants on the sorbent performance. In Task 6, we designed the CO_2 separation system and the PSA cycle sequence that will best meet the needs of the proposed application. In Task 7, we performed a detailed engineering design for the construction of the prototype test unit that was installed at the NCCC and the Wabash River IGCC Power Plant. In Task 8, we carried out a detailed system analysis and economic assessment of the new technology and compared against other currently available technologies such as cold gas CO_2 capture with Selexol^{TM} solvent scrubbing technology. In Task 9, we fabricated the prototype unit and tested it in-house prior to shipment to qualify it for use at the demonstration sites. In Task 10, we worked with Southern Company and Phillips 66 to install the prototype unit at the demonstration sites and carry out 3-week test campaigns. Task 11 involved all the project management activities and reporting.

4. Results

4.1 Task 1. Sorbent Optimization and Characterization

In this task we produced a variety of CO_2 sorbent formulations, optimizing the sorbent by varying its chemical composition and physical properties. In the sorbent synthesis work, we used raw materials that are widely available in bulk at low cost. The production process used only scalable production equipment and techniques.

Figure 5. Micromeritics Gemini Unit.

4.1.1 Sample Optimization

TDA prepared more than 20 different sorbent formulations based on our proprietary synthesis method. During these preparations we varied the amount and type of precursors, promoters and raw materials to introduce the CO_2 selective surface functional groups. We also controlled the pore size and surface area of these formulations by changing the activation temperature and duration.

Table 2. Physical properties of various sorbent samples prepared.

Sample	**BET Surface Area (m^2/g)**	**DFT Total Pore Volume (cc/g)**	**DFT Micropore Volume (cc/g)**
AMS-19 (900°C)	401	0.17	0.17
AMS-19 (1000°C)	505	0.21	0.21
AMS-84 (900°C)	879	0.36	0.35
AMS-92 (900°C)	556	0.24	0.22
AMS-93 (900°C)	954	0.43	0.34
AMS-93 (550°C)	875	0.39	0.33
AMS-93 (500°C)	732	0.33	0.28
AMS-94 (900°C)	707	0.29	0.29
AMS-97 (900°C)	900	0.39	0.34

After making the sorbents, we measured their BET surface area and pore volume in our Micromeritics Gemini surface area analyzer (Figure 5). Table 2 shows the BET surface area and pore volume for selected formulations prepared in this project. Figure 6 shows the pore size distribution for the AMS-19 sorbent calculated using density functional theory (DFT) from the nitrogen adsorption isotherm at 77K. Most of the sorbent porosity is in the meso-range pores (20 to 100 Å), these relatively large pores that eliminate the pore diffusion-based mass transport limitations and allow rapid cycling of the sorbent. The mesoporosity of the sorbent is also beneficial since it allows us to introduce other functional groups for capturing trace contaminants in the synthesis gas. The sorbent had a crush strength of 1 lbf/mm, which is comparable to commercial carbon sorbents that have been widely used in various chemical processes.

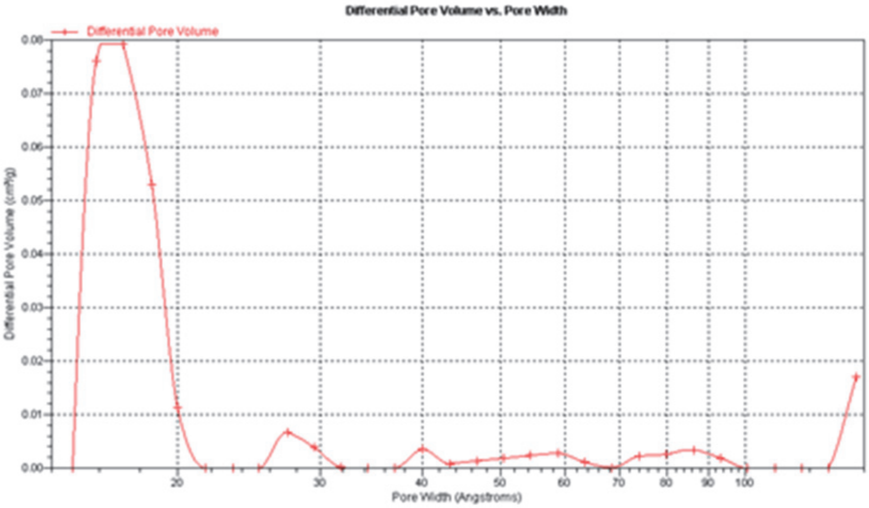


Figure 6. DFT pore size distributions measured from nitrogen isotherms for AMS-19.

4.1.2 Thermo-Gravimetric Analysis

The performance of these new sorbent formulations was initially evaluated in a Thermo-Gravimetric Analyzer (Shimadzu TGA-50 as shown in Figure 7). In these tests, we quickly measured the CO_2 adsorption capacity of these formulations at two different temperatures (60 and 180°C). We cycled the sorbent by flowing pure CO_2 and N_2 to simulate adsorption and regeneration while keeping the adsorption temperature constant (at a CO_2 and N_2 total pressure of about 12 psia).

Figure 7. Shimadzu TGA-50.

In the final application the CO_2 adsorption partial pressure would be significantly higher. Hence, these low pressure tests were only used as a screening tool. Figure 8 shows the results from the CO_2/N_2 cycling for AMS-19 sorbent at 180°C. Similar results were obtained for other samples. The CO_2 adsorption capacity from the TGA screening experiments for selected sorbent samples are provided in Figure 9.

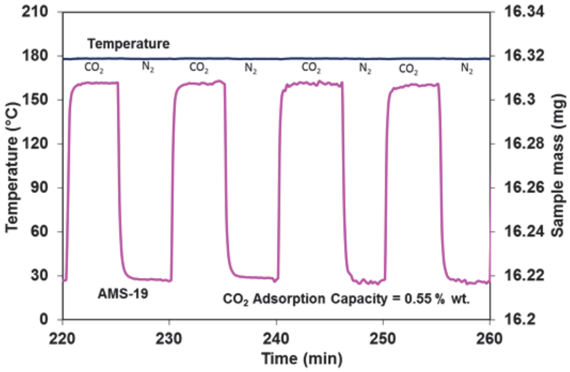


Figure 8. CO_2/N_2 adsorption/desorption cycling results obtained in TGA for AMS-19 (1000°C) at 180°C.

We found that AMS-19 provides the best performance at 60°C, whereas AMS-93 (500°C, indicates the firing temperature at which the sorbent is made) outperformed all other samples at 180°C. The samples that showed desirable CO_2 absorption capacity were further selected for screening in a bench-scale fixed bed flow reactor. The sorbent performance was evaluated under representative conditions using simulated gas mixtures.

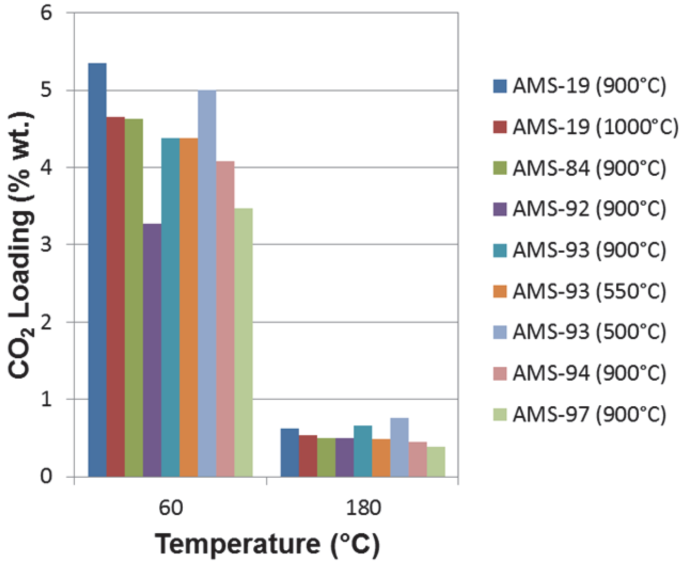


Figure 9. CO_2 adsorption capacity from sorbent screening with TGA at 60°C and 180°C.

4.1.3 Chemical Analysis

We sent several samples for elemental analysis to Hazen Research (Golden, CO). In Figure 10 and Figure 11 highlights the nitrogen and oxygen content of the selected sorbent samples as a function of the CO_2 loading at 180°C. In agreement with the theory and despite some scatter, the higher the nitrogen and oxygen content in the sorbent resulted in higher CO_2 capacity.

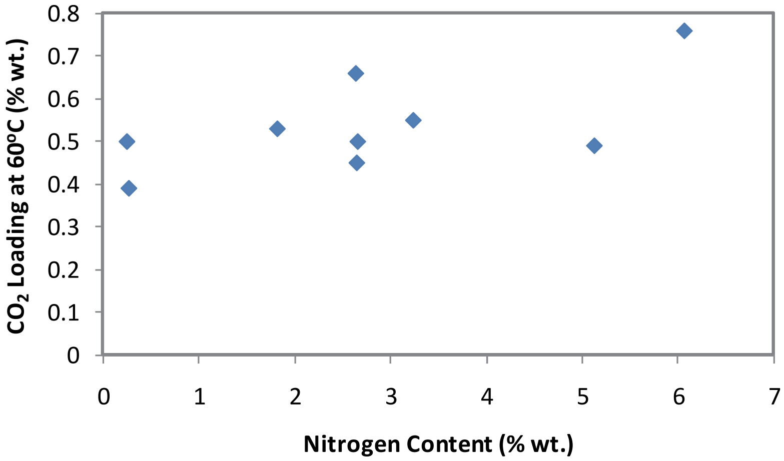


Figure 10. CO_2 loading versus nitrogen content at 180°C.

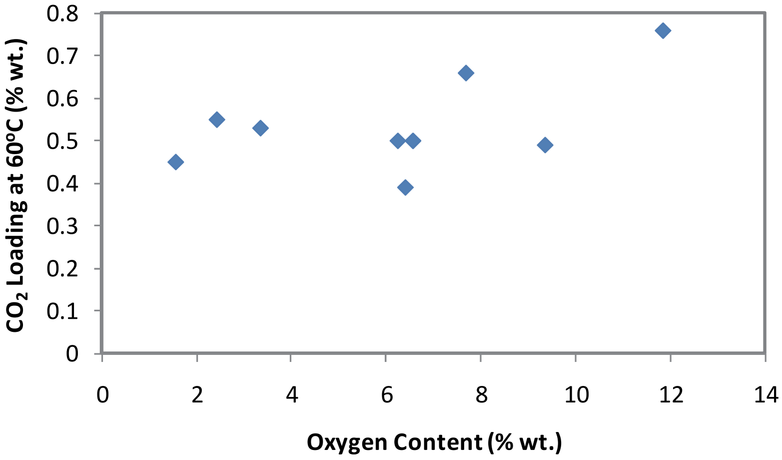


Figure 11. CO_2 loading versus oxygen content at 180°C.

4.1.4 Rate of Adsorption

Figure 12 shows the comparison of adsorption and desorption curves at various temperatures as measured in the TGA at different temperatures. At higher temperatures the rate of adsorption is higher and the sorbent reaches the equilibrium faster. The desorption curves are scattered and do not show any trends.

In general, physical adsorption and desorption of CO_2 in sorbent structures (similar to that of TDA's) usually follow the Langmuir kinetic (rate) model due to the slit cavity structures of their pores. The slit-molecule interaction potential controls the rate of adsorption and desorption of CO_2 molecules into the pores by controlling resistance for adsorption due to the constrictions present in pore mouths).

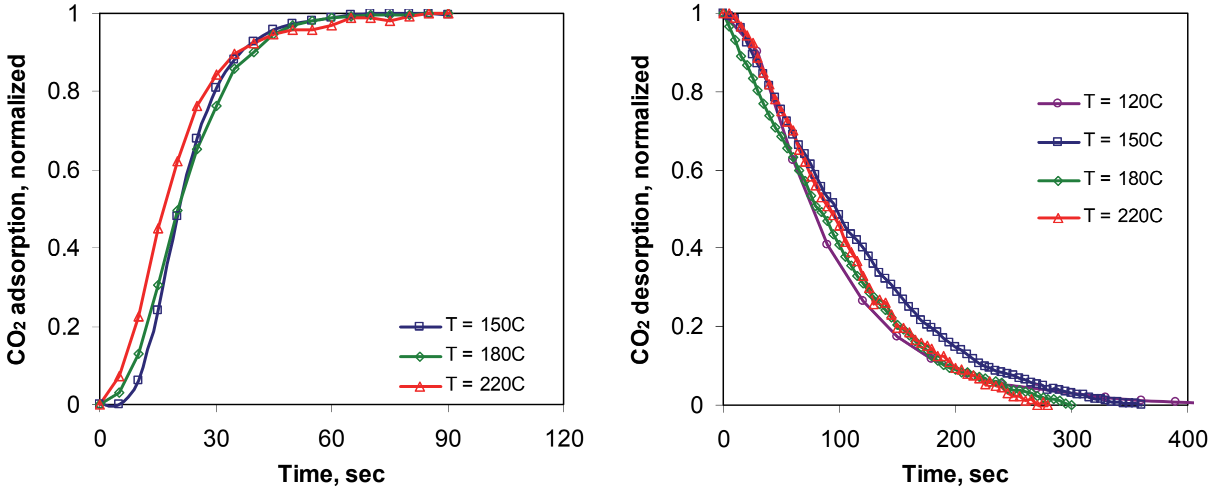


Figure 12. CO_2 adsorption and desorption as a function of time at various temperatures.

Langmuir rate model (also known as barrier resistance model)

$$\frac{dq}{dt} = k_b(q^* - q) \qquad (1)$$

The solution to barrier resistance model for uptake in a TGA is given as

Normalized adsorption, $\frac{m_t}{m_\infty} = 1 - \exp(-k_b t)$ (2)

Figure 13 shows the normalized adsorption and desorption curves along with fitted the barrier resistance model and the rate constant (k_b). The rate constant for adsorption and desorption of CO_2 as measured in TGA at 12 psia partial pressure is 0.065 s^{-1} and 0.0115 s^{-1} respectively. This rate of adsorption and the rate constants would increase with increase in CO_2 partial pressure. The rate constant is plotted against reciprocal of temperature in Figure 14, which follows Arrhenius type temperature dependence shown in equation (3) and the activation energy for CO_2 adsorption is estimated to be 1 kcal/mol.

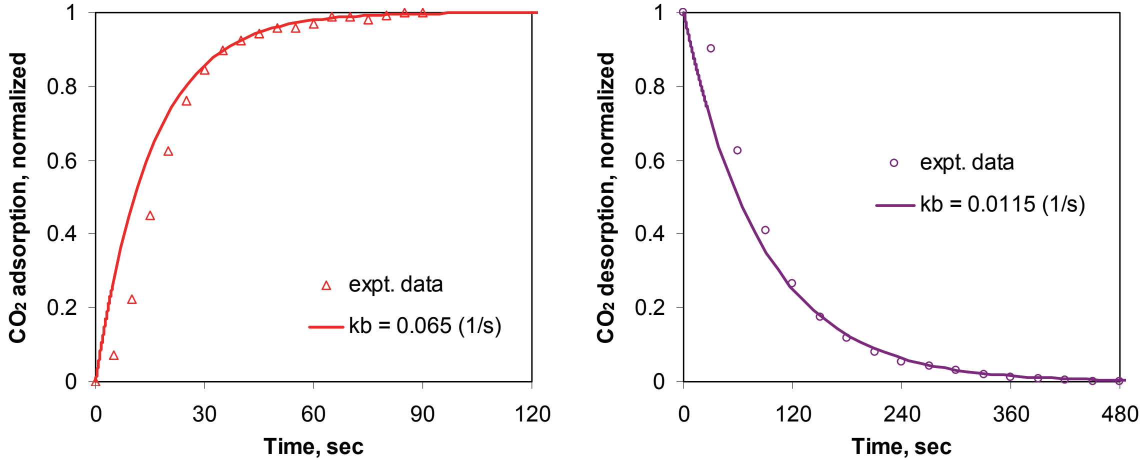


Figure 13. CO_2 adsorption and desorption curves at 220°C fitted with Langmuir rate model.

$$k_b = k_{b0} \exp\left(-\frac{E}{R_g T}\right) \quad (3)$$

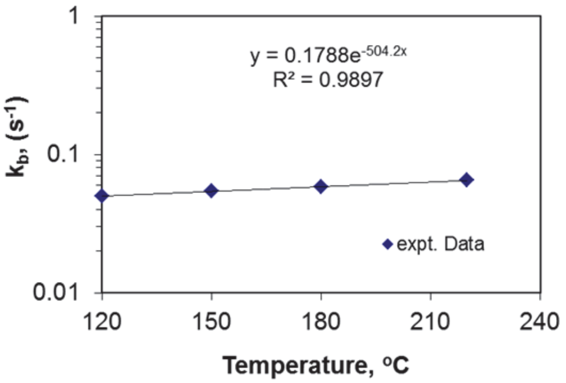


Figure 14. Activation energy plot for CO_2 adsorption on TDA's Sorbent.

4.1.5 Heat of Adsorption

We estimated the heat of adsorption of CO_2 over the sorbent from the CO_2 capacity measurements at different temperatures. Assuming the isotherm is linear and in the Henry's Law regime at the low partial pressures of CO_2 used in the measurements, we estimated the heat of adsorption using the following equation:

$$q = K_0 \exp\left(\frac{-\Delta H}{R_g T}\right) P_{CO2} \quad (4)$$

Figure 15 shows the plot of temperature against CO_2 adsorption capacity at a partial pressure of about 12 psia as measured in the TGA, fitted with equation (4). The heat of adsorption of CO_2 on TDA's sorbent is estimated to be 5.4 kcal/mol.

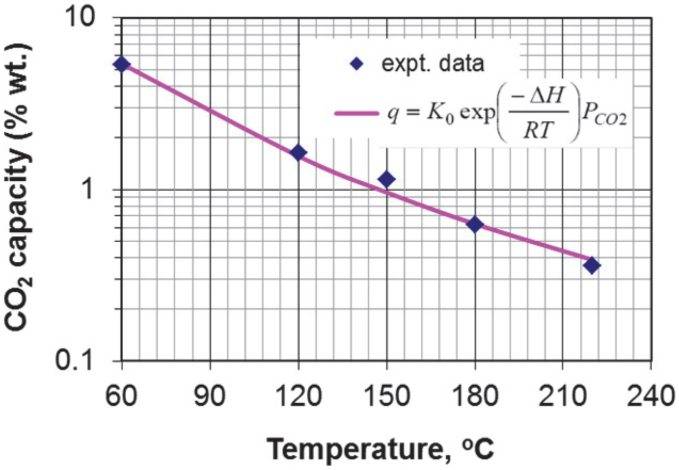


Figure 15. Heat of adsorption plot for CO_2 adsorption on TDA's Sorbent from TGA measurements.

We also measured the heat of adsorption using a differential scanning calorimeter (Shimadzu DSC-50, Figure 16). The Shimadzu DSC-50 is a heat flux differential scanning calorimeter (DSC). In the DSC-50 the sample and reference are heated and/or cooled or kept at a constant temperature in the same furnace with a controlled atmosphere. As the sample is heated or cooled or the atmosphere is changed, it will go through phase transitions (i.e., adsorb or desorb). During these transitions, the sample will either absorb or radiate heat. This heat change is characterized by a temperature change in the sample which is detected by a thermocouple and compared with the relative temperature of the reference cell ($T_s - T_r = \Delta T$). The ΔT is proportional to changes in the calorimetric state of the sample and the resultant signal is reported in mW. We used this temperature change in the DSC to measure the heat of adsorption and desorption of CO_2 normalized on a per g sample basis. Figure 17 shows a typical data in the DSC when cycled between CO_2 and N_2. Testing in parallel in a Thermo Gravimetric Analyzer (TGA) under similar test conditions provided the amount of CO_2 adsorbed and or desorbed per unit mass of the sorbent. Using the two measurements the heat of adsorption or desorption is normalized on a per mol of CO_2 adsorbed or desorbed basis.

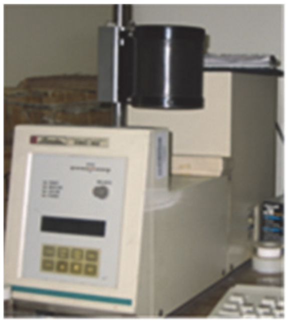


Figure 16. Shimadzu DSC-50.

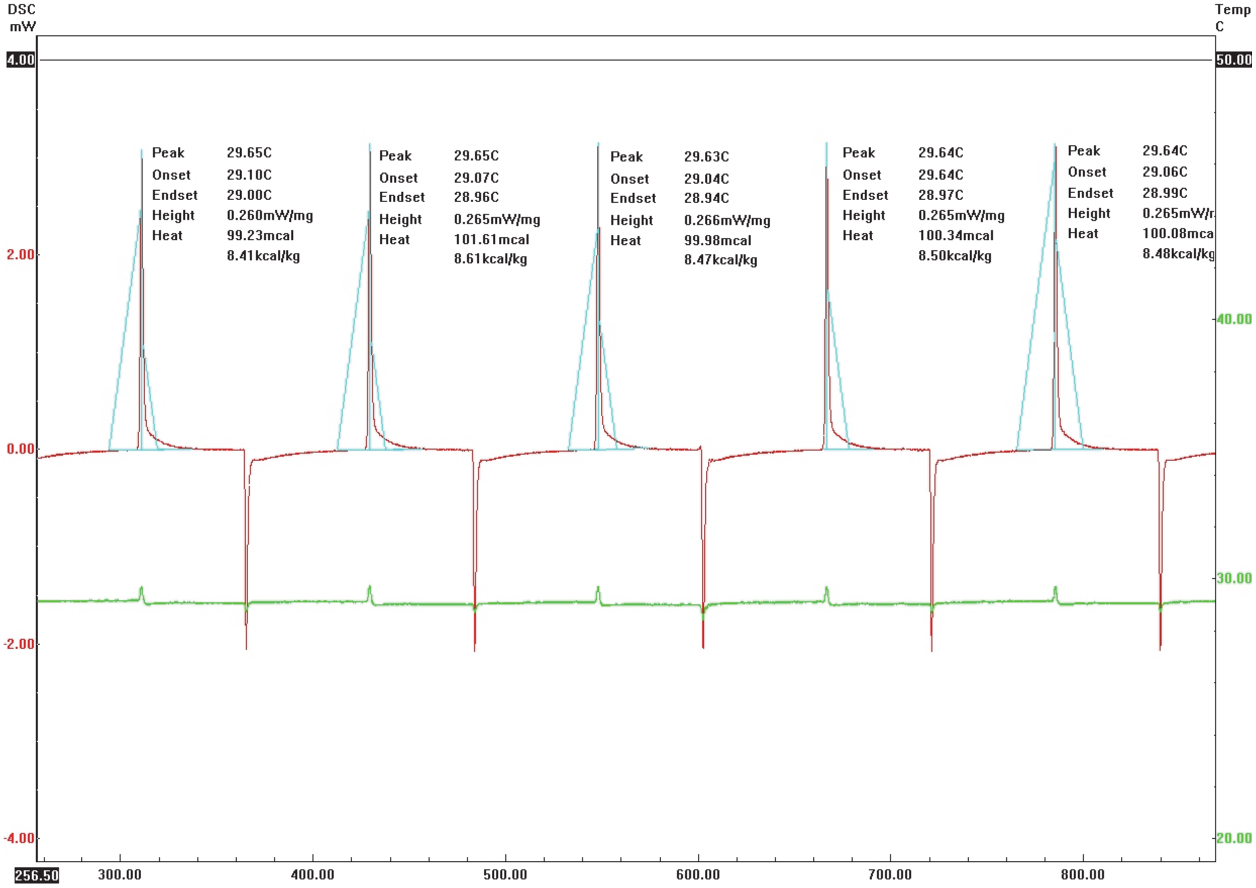


Figure 17. DSC data showing the temperature changes in the sorbent sample during adsorption and desorption.

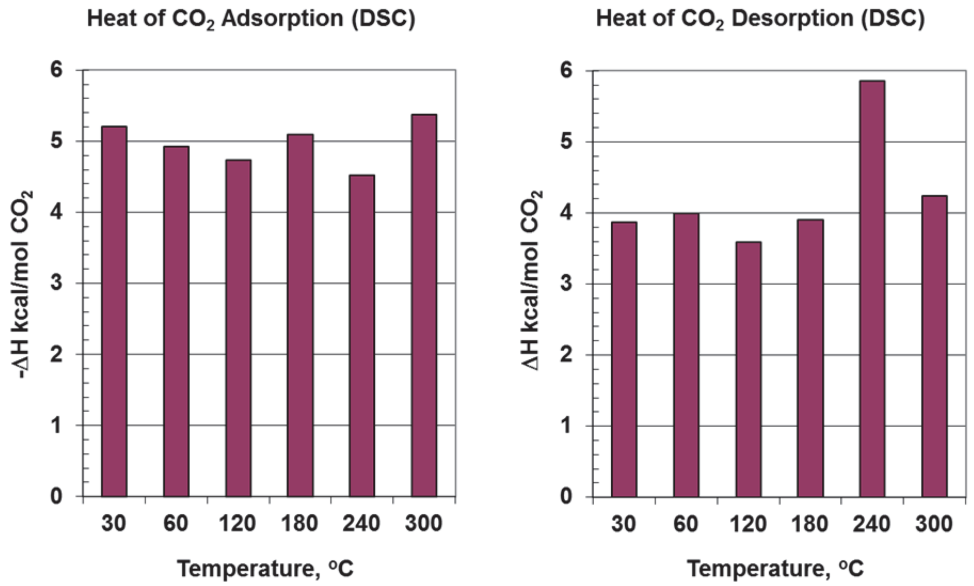


Figure 18. Heat of adsorption and desorption of CO_2 on TDA's sorbent as measured in the combined DSC-TGA experiments.

From the combined DSC-TGA measurements the heat of adsorption ($-\Delta H_{ads}$) of CO_2 on TDA's sorbent is 4.9 ±0.4 kcal/mol and the heat of desorption (ΔH_{des}) of CO_2 on TDA's sorbent is 4.3 ±0.8 kcal/mol. The heat of adsorption and desorption as measured in the DSC-TGA experiments at various adsorption and desorption temperatures are shown in Figure 18.

4.1.6 Comparison Against Commercial Sorbents

We also benchmarked the performance of our sorbent (AMS-19) against the commercial carbon sorbents from Calgon and Norit (Figure 19). TDA's sorbent achieved significantly higher CO_2 capacity of about 55% and 30% higher than Calgon BL sample at 60 and 180°C respectively.

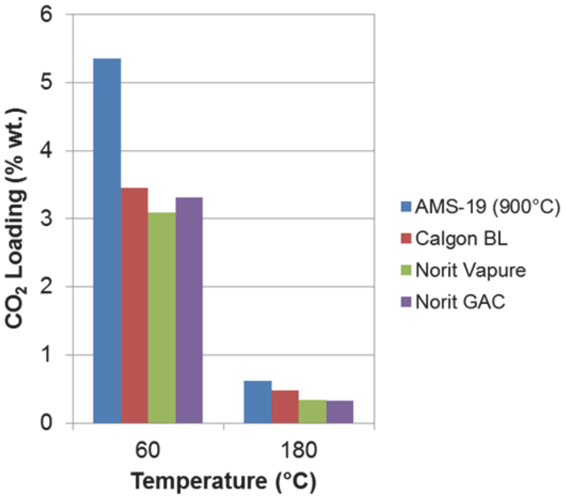


Figure 19. Comparison of CO_2 adsorption capacity of TDA's sorbent against commercial sorbents from sorbent screening with TGA at 60 and 180°C.

4.2 Task 2. Evaluation of the Sorbent at Bench-scale

In this task, we evaluated selected sorbent samples from the TGA tests. Our focus in these evaluations was to screen the samples then select the best formulation and generate data to develop CO_2 adsorption isotherms at different temperatures (i.e., we measured the CO_2 adsorption capacity as a function of CO_2 partial pressure and temperature). As part of these tests, the selected AMS-19 sorbent was subjected to more than 450 cycles under isobaric and isothermal conditions. To provide experimental convenience, during the early tests we used mixtures of CO_2, N_2 and H_2O for adsorption and N_2 and mixtures of N_2 and H_2O for the regeneration.

4.2.1 Bench-scale System Description

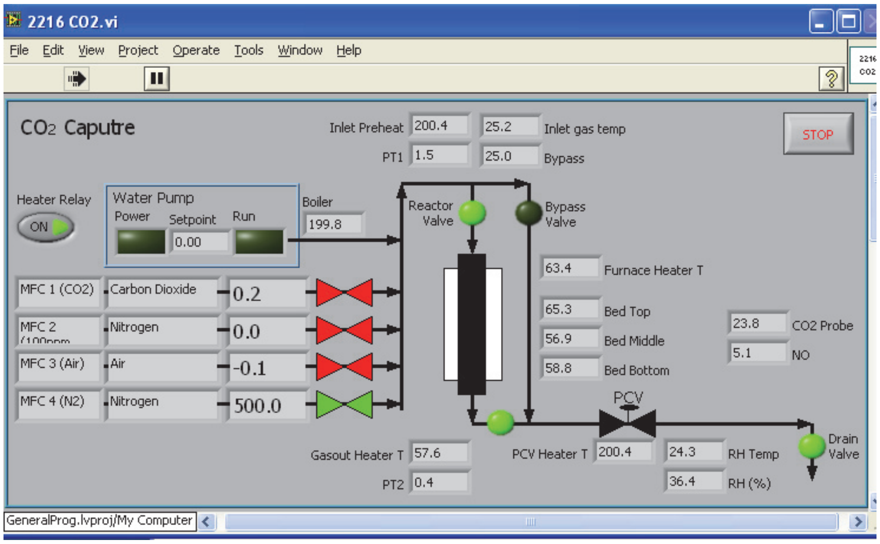
We modified an existing testing apparatus that is capable of operation at pressures up to 900 psig and temperatures up to 300°C to evaluate the sorbent's performance under representative conditions (Figure 20). The apparatus was built at TDA for the specific purpose of measuring the activity of sorbents (with minimal dead volume in the reactor ullage space and in the manifolds). In this system, the desired gas mixtures were directed into a fixed-bed micro-reactor that houses granules or pellets of the sorbent. All gas flows are controlled with electronic mass flow controllers. A Waters pump is used to introduce water. After mixing in a manifold, the feed gas mixture is then directed into the reactor. A valve system allows the gases to bypass the reactor and flow directly to the analytical system for accurate measurement of the feed gas composition.

Figure 20. Bench-scale sorbent test apparatus.

In typical tests, the sorbent reactor is loaded with 100 mL of 4-20 mesh size sorbent. The reactor has three thermocouple ports to monitor the sorbent bed temperature. A back pressure regulator is used to control the adsorption pressure (as needed). After exiting the reactor, the

CO_2 content of the stream are monitored by an on-line NOVA Multi-gas Analyzer and Vaisala CO_2 and humidity probes. Continuous analysis of CO_2 allows us to monitor breakthrough gas concentrations and to measure a total CO_2 adsorption capacity.

The apparatus is fully automated using a LabVIEW control system and can run without an operator for long periods of time, including overnight. The LabVIEW system controls the test conditions, logs the analytical data, and also safely shut down the apparatus in case of a malfunction. Figure 21 shows the screen shot of the LabVIEW control system.



Inlet Preheat 200.4
PT1 1.5
Inlet gas temp 25.2
Bypass 25.0
Setpoint 0.00
Boiler 199.8
Furnace Heater T 63.4
Bed Top 65.3
Bed Middle 56.9
Bed Bottom 58.8
CO2 Probe 23.8
NO 5.1
MFC 1 (CO2) Carbon Dioxide 0.2
MFC 2 Nitrogen 0.0
MFC 3 (Air) Air -0.1
MFC 4 (N2) Nitrogen 500.0
Gasout Heater T 57.6
PT2 0.4
PCV Heater T 200.4
RH Temp 24.3
RH (%) 36.4

Figure 21. LabVIEW screen shot of the test apparatus.

A typical CO_2 breakthrough curve obtained in these tests is shown in Figure 22 along with the breakthrough for reactor loaded with quartz beads (marked as blank). The blank curve indicates the dead volume in the system. The CO_2 adsorption capacity is calculated by integrating the area between the blank and the sorbent breakthrough curves.

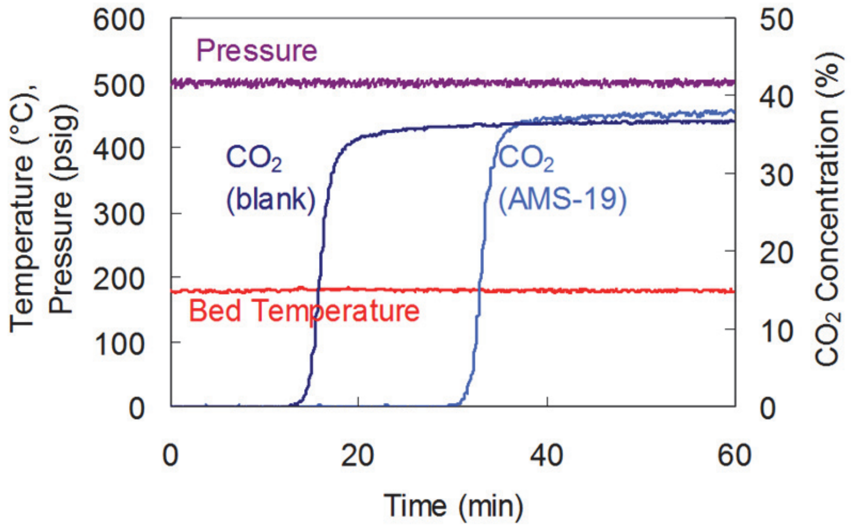


Figure 22. Typical CO_2 breakthrough in the bench-scale system. AMS-19 at 240°C and 500 psig, adsorption gas: 40% CO_2 in N_2, desorption gas: N_2.

4.2.2 Screening Tests

We carried out tests with both AMS-19 (900°C) and AMS-93 (500°C) sorbents in flow experiments at two different adsorption temperatures and CO_2 partial pressures. Multiple adsorption/ desorption cycles at each of these test conditions to ensure steady state sorbent operation. Table 3 summarizes the test conditions along with the CO_2 capacity. The AMS-19 sorbent achieved significantly higher capacity at both the temperatures (180 and 240°C) and CO_2 partial pressures (150 and 250 psig). Hence, we selected the AMS-19 sorbent for further parametric tests.

Table 3. Summary of sorbent screening in bench-scale system.

Adsorption	Desorption	AMS-93	AMS-19
240ºC, 30% CO2 70% N2, 500 psig	240ºC, 100% N2, 500 psig	3.55	4.20
240ºC, 50% CO2 50% N2, 500 psig	240ºC, 100% N2, 500 psig	5.22	6.35
180ºC, 30% CO2 70% N2, 500 psig	180ºC, 100% N2, 500 psig	3.98	6.82
180ºC, 50% CO2 50% N2, 500 psig	180ºC, 100% N2, 500 psig	6.14	9.89

4.2.3 Cycling/ Parametric Tests

In the parametric tests, we extensively tested the AMS-19 sorbent formulation under different adsorption temperatures and CO_2 partial pressures. The results are provided in Figure 23. The sorbent maintained an average 6% wt. CO_2 capacity at CO_2 partial pressures of 120 psig. As expected, higher CO_2 capacities were observed as we lower the adsorption temperature and increase the CO_2 partial pressure. In these tests, we also showed that the H_2O has little or no effect on the CO_2 adsorption capacity of the sorbent. The data highlighted in Figure 23 (data circled in) shows that the CO_2 capacity remained unchanged when 15% vol. H_2O was added to the adsorption gas. The test results also suggest that the sorbent maintains its capacity for more than 650 cycles with no signs of degradation due to cycling and can achieve capacities in excess of 20% wt. CO_2 at higher CO_2 partial pressures.

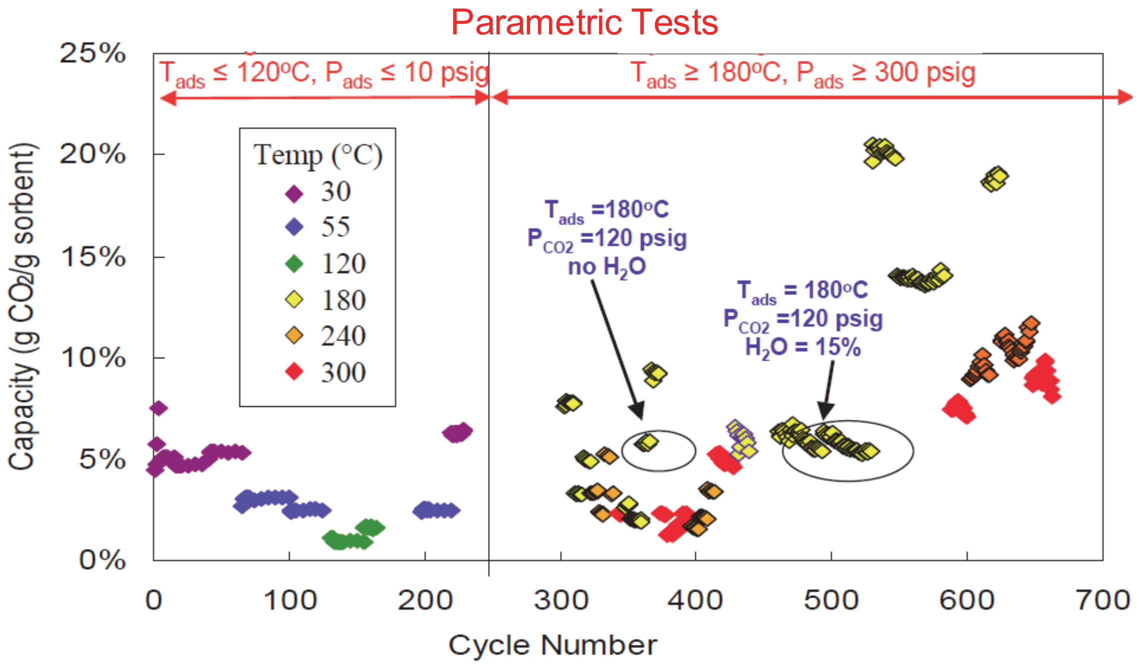


Figure 23. Bench-scale sorbent test results under CO_2/N_2.

4.2.4 Langmuir Freundlich Isotherm

The experimental data from the bench-scale tests were fitted with Langmuir Freundlich adsorption isotherm model. Langmuir Freundlich Isotherm Model equation is shown below:

$$q = \frac{q_s B P^n}{1 + B P^n} \tag{5}$$

Where

$$q_s = k_1 e^{k_2/T}; \quad B = k_3 e^{k_4/T}; \quad n = k_5 e^{k_6/T} \tag{6}$$

Table 4 shows the fitted the Langmuir Freundlich isotherm model parameters. Figure 24 shows the CO_2 isotherms on AMS-19 at different temperatures with the fitted Langmuir-Freundlich isotherms.

Table 4. Langmuir Freundlich isotherm model parameters.

q (mol CO_2/kg); P (psia); T (K)

k1	58.05	k4	47.07
k2	46.55	k5	0.59
k3	2.2E-04	k6	201.46

4.2.5 Isosteric Heat of Adsorption

Isosteric Heat of adsorption was calculated at different sorbent loadings from the adsorption equilibrium data in Figure 24 using Clasius-Clapeyron equation.

$$\frac{\Delta H_{ads}}{R_g} = \frac{d(\ln P)}{d\left(\frac{1}{T}\right)} \tag{7}$$

$$\ln\left(\frac{P_1}{P_2}\right) = \frac{\Delta H_{ads}}{R_g}\left(\frac{1}{T_2} - \frac{1}{T_1}\right) \tag{8}$$

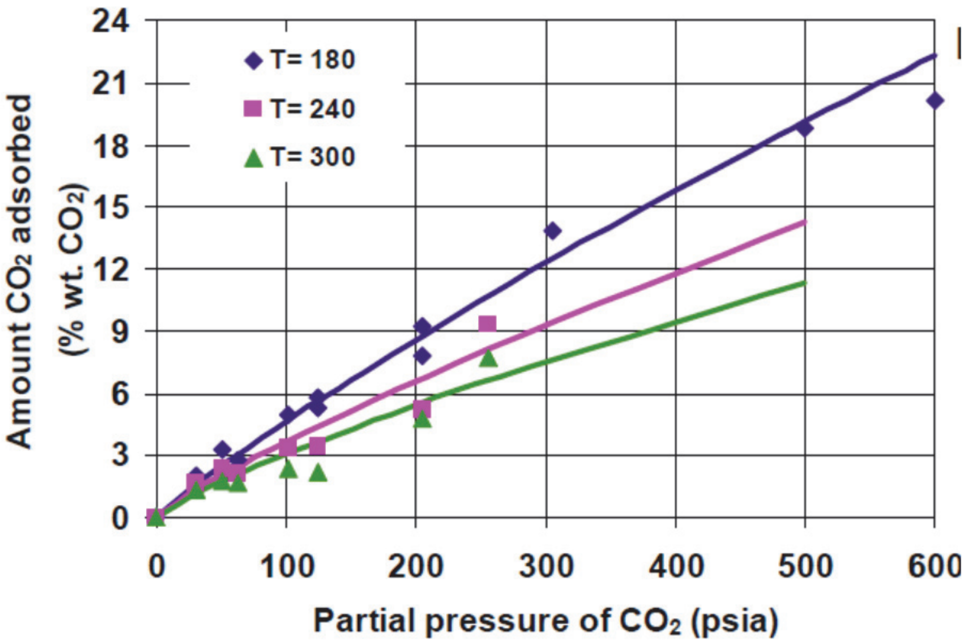


Figure 24. CO_2 Isotherms at different temperatures (T= 180, 240 and 300°C) and the fitted Langmuir-Freundlich isotherm parameters.

Figure 25 shows the isosteric heat of adsorption from Clasius–Calpeyron Equation as a function of sorbent loading (i.e., CO_2 surface coverage). The isosteric heat of adsorption is estimated to be between 4-6 kcal/mol, which is similar to those measured in the DSC-TGA tests (4.9 ±0.4 kcal/mol). The isosteric heat of adsorption decreases at higher sorbent loading since the initially more energetic surface adsorption sites are filled before less energetic sites start adsorbing CO_2.

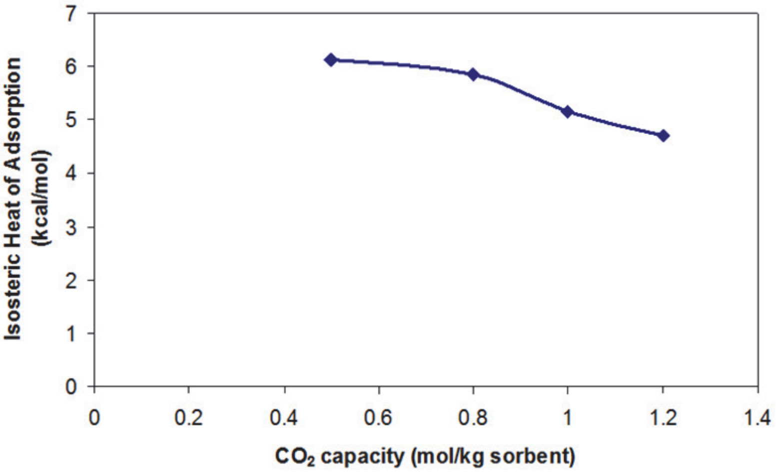


Figure 25. Isosteric heat of adsorption from Clasius-Clapeyron Equation at different sorbent loadings.

4.3 Task 3. Process Design and Modeling

In this task, TDA, in collaboration with UCI performed a detailed process design and modeling using Aspen-Plus^{TM} simulation software to calculate material and energy balances. TDA prepared a preliminary system design for the CO_2 control system and conducted an initial process evaluation. In this task, UCI completed the preliminary process design and modeling of the IGCC power plant integrated with TDA's warm gas cleanup unit. To calibrate the model, UCI also conducted a similar analysis for the Selexol^{TM}-based solvent scrubbing system, referred to as Cold Gas Cleanup Case.

Consistency has been maintained between the IGCC case developed under this study utilizing the TDA's regenerable sorbent CO_2 capture process (Warm Gas Cleanup Case) with the Phillips 66 (E-Gas^{TM}) gasifier based IGCC with CO_2 capture utilizing current state-of-the-art Selexol^{TM} based syngas cleanup and CO_2 capture technology (Cold Gas Cleanup Case) that makes up Case 4 in the DOE NETL study report DOE/NETL-2007/1281 titled, "Cost and Performance Baseline for Fossil Energy Plants," dated August 2007.

The analysis includes three simulations:

- E-Gas^{TM} based IGCC plant with Selexol-based CO_2 capture
    - Calibration Case
    - Compare/validate model results with prior DOE/NETL analysis
- E-Gas^{TM} based IGCC plant with Selexol - 90% CO_2 capture
- E-Gas^{TM} based IGCC plant with TDA's CO_2 capture system

UCI first modeled the Calibration Case in Aspen Plus® and the overall thermal performance of the plant was compared to that of the DOE/NETL study Case 4 in order to validate the model predictions. This model was then modified to reduce the CH_4 content in the gasifier effluent in order to increase the carbon capture from 88.2% for the Calibration Case to 90% (as required by the FOA). This model was further modified to include the Warm Gas Cleanup while the CH_4 content in the gasifier effluent was maintained at the same level (as the Cold Gas Cleanup Case), all the while maintaining the 90% carbon capture.

4.3.1 Cold Gas Cleanup

The IGCC plant employing the cold gas cleanup and CO_2 capture technology consists of the following plant subsystems:

- Air Separation Unit (ASU)
- Coal Feed Preparation
- Gasification (based on Conoco Phillips technology)
- High Temperature Syngas Cooling and Scrubbing
- Sour Shifting and Cold Gas Heat Recovery
- Syngas Desulfurization and Decarbonization using a two-stage Selexol™ process
- Claus Sulfur Recovery and Tail Gas Hydrogenation followed by Recycle to the Gasifier
- CO_2 Dehydration and Pressurization
- Gas Turbine (based on GE F class technology)
- Heat Recovery Steam Generator Unit
- Reheat Steam Cycle

The plant also has the necessary utilities (e.g. cooling water supply, make-up water treatment, plant and instrument air) to support the process units. A detailed process description of this case may be found in the previously referenced DOE/NETL report.

4.3.2 Warm Gas Cleanup

The IGCC plant employing the sorbent CO_2 capture consists of the following plant subsystems:

- Air Separation Unit (ASU)
- Coal Feed Preparation
- Gasification (based on Conoco Phillips technology)
- High Temperature Syngas Cooling and Scrubbing
- Warm Gas Cleanup similar to RTI’s process including Acid (H_2SO_4) Unit
- Sour Shifting
- Regenerable Sorbent CO_2 Capture (based on TDA technology)
- CO_2 Purification and Pressurization
- Gas Turbine (based on GE F class technology)
- Heat Recovery Steam Generator Unit
- Reheat Steam Cycle

The overall configuration of the plant is shown in Figure 26. Plant subsystems that are different from the Cold Gas Cleanup cases are described below:

1. About 28% of the raw syngas exiting the scrubber is cooled in a heat exchanger while recovering the heat for vacuum condensate/makeup BFW heating followed by trim cooling against cooling water before it is compressed and recycled back to the gasifier system as quench gas.

2. The remainder of the scrubbed gas is preheated to a temperature of 260°C in a feed/effluent exchanger and supplied to a warm gas cleanup unit similar to RTI’s process for removal of sulfur compounds utilizing a Zn titanate adsorbent in a fluidized bed. The performance of this unit as well as the production of H_2SO_4 from the SO_2 in the regenerator off-gas was developed utilizing information available in the public domain. The regenerator off-gas after particulate removal is depressurized by expansion in a power recovery turbine before feeding it to the H_2SO_4 unit. The on-site ASU provides the small amount of O_2 as required by the H_2SO_4 unit in addition to the gasifier. The hot synthesis gas leaving the desulfurizer is cooled to a temperature of 233°C in the feed/effluent exchanger where the desulfurizer feed gas is preheated.

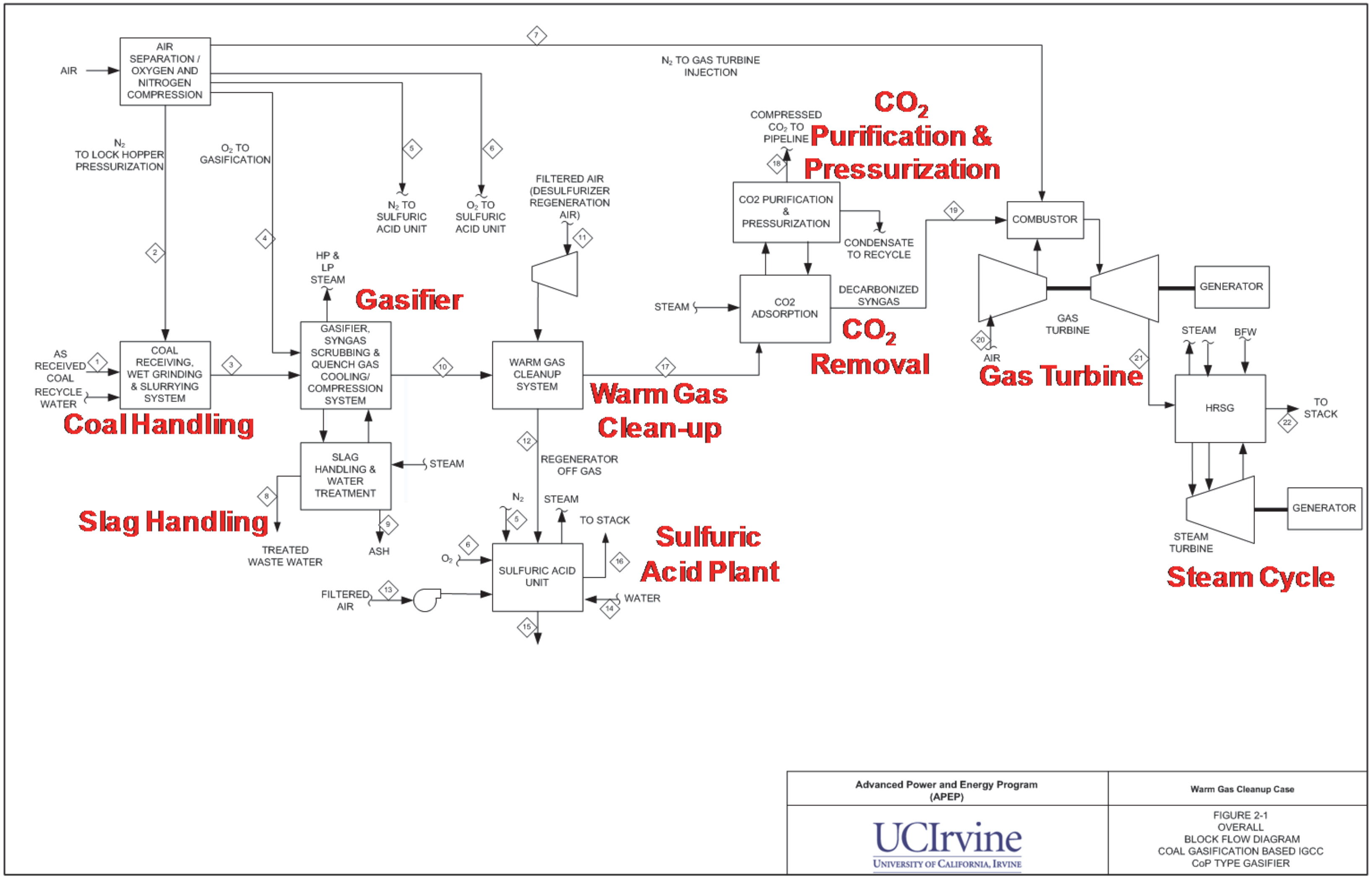


Figure 26. TDA's System with IGCC.

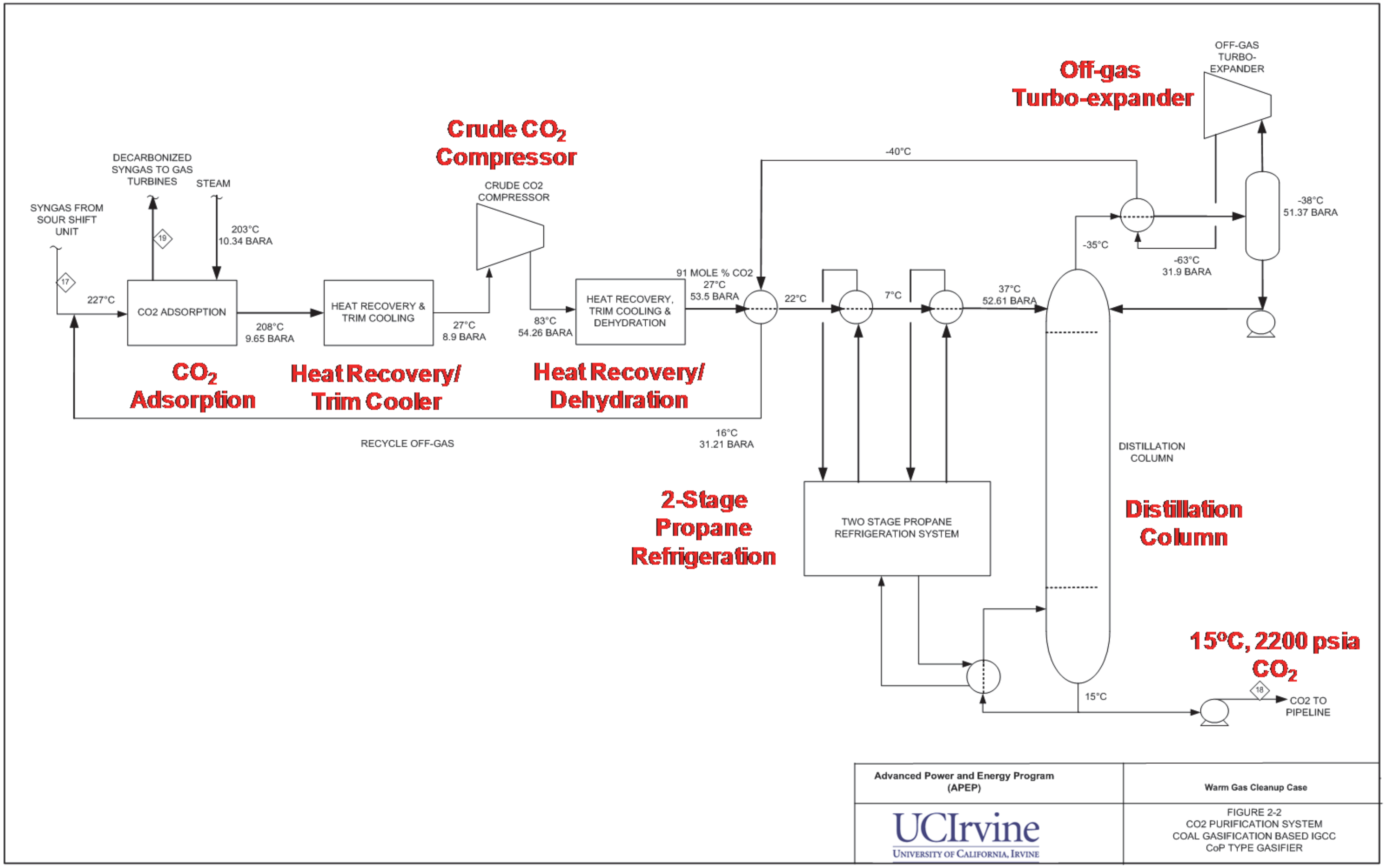


Figure 27. TDA's System with CO_2 purification and compression sub systems.

3. Next, the desulfurized synthesis gas is treated to remove trace components such as Hg in a fixed bed sorbent scrubber consisting of a proprietary adsorbent developed by TDA (In a previous SBIR Phase II project DOE DE-FG02-08ER85211). Some of the NH_3 and HCN are also captured in the fixed bed adsorbent. The clean gas is then treated in a sour shift unit similar to the Cold Gas Cleanup cases consisting of three adiabatic beds in series with intercooling where intermediate pressure (IP) and medium pressure (MP) steam is generated. Steam required by the shift unit is supplied as attemperated steam extracted from the steam cycle.

4. The shifted synthesis gas leaving the last shift reactor at a temperature of 234°C is combined with recycle gas exiting the CO_2 purification unit and then fed to the TDA fixed bed adsorption unit for decarbonizing the syngas before it is combusted in the gas turbines. More than 90% of the syngas enters this decarbonizing unit where 95% of the CO_2 entering with the syngas is separated on a per-pass basis. Remainder of the syngas is utilized in the bed re-pressurization cycle, the adsorbent bed being regenerated at a pressure of 10.34 barA (150 psia). Regeneration is accomplished utilizing 1.31 moles steam per mole CO_2. The mixture consisting of CO_2, steam and residual syngas at a temperature of 208°C is then cooled in a series of heat exchangers while generating low pressure (LP) steam, vacuum condensate / makeup BFW heating and finally trim cooling against cooling water before it is compressed, cooled while recovering bulk of the heat for vacuum condensate / makeup BFW heating, dehydrated and fed to the cryogenic CO_2 purification unit (see Figure 27). The distillation column operates at a nominal pressure of 52 barA and its primary function is to produce a CO_2 product that meets the pipeline specifications while minimizing CH_4 losses (dissolved in the liquid CO_2 bottoms stream leaving the column). The bottoms stream is pressurized to the pipeline specification. The feed to the column is cooled in a series of heat exchangers to a final temperature of 37°C against cold process streams as well as refrigerated liquid propane.

5. The decarbonized synthesis gas leaving the adsorption unit at a temperature of 265°C with its accompanying unreacted steam is supplied to the gas turbines along with pressurized N_2 from the ASU. The amount of N_2 added to the gas turbine is significantly lower than that in the Cold Gas Cleanup cases, however, due to the large amount of water vapor present in the syngas. The combined cycle design is similar to the design in the Cold Gas Cleanup cases consisting of a reheat steam cycle.

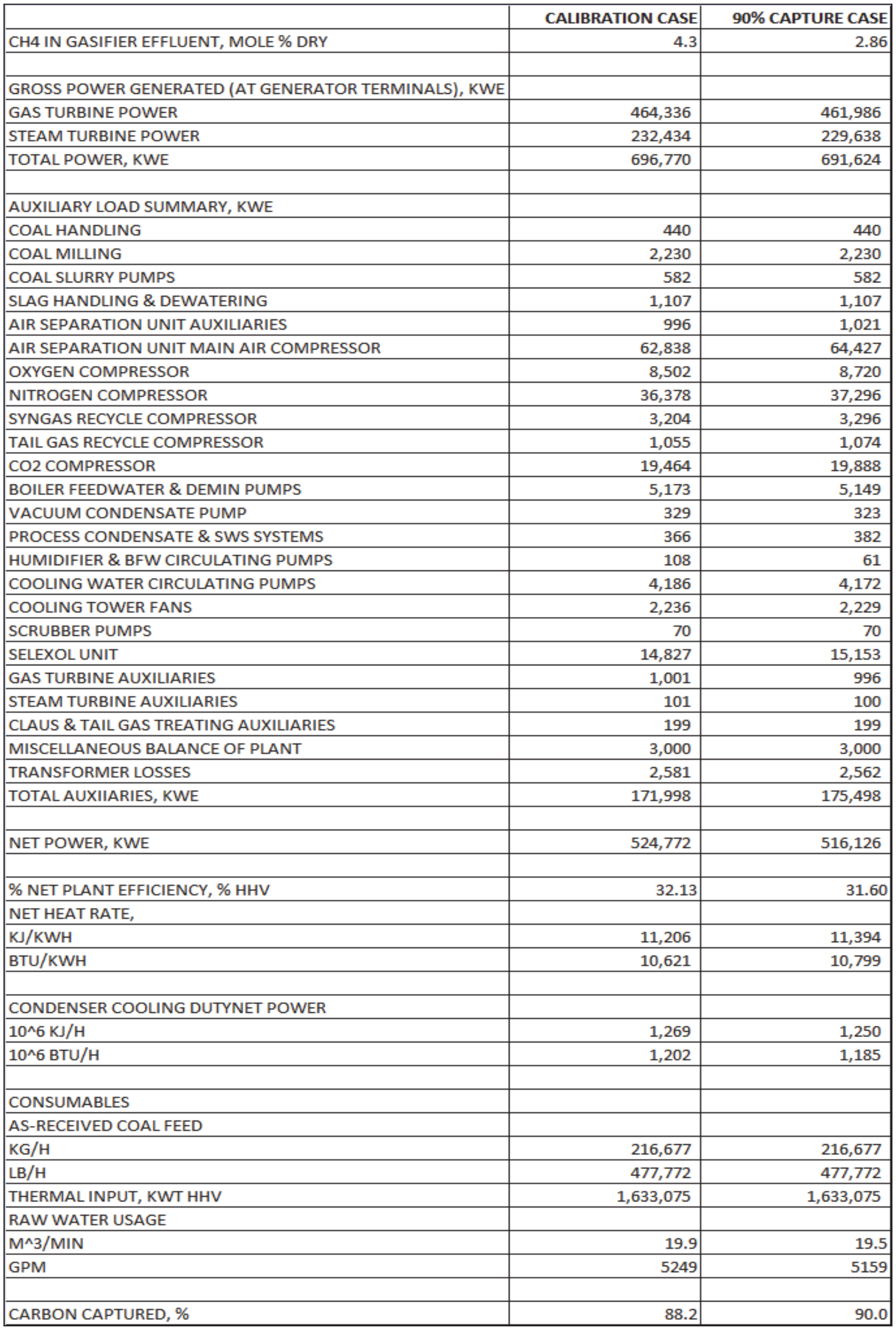
4.3.3 Preliminary System Analysis Results

The plant performance of the Calibration Case (which was first modeled in Aspen Plus® to compare its overall plant thermal performance with the DOE/NETL study Case 4 in order to validate the Aspen Plus® IGCC system model) is summarized in Table 5. The calculated net thermal efficiency of this case is slightly higher at 32.13% versus 31.7% (both on a coal HHV basis) as reported in the DOE/NETL study Case 4. The primary reason for the higher efficiency is that the CO_2 pressurization was accomplished by a combination of gas compression and liquid CO_2 pumping which is more efficient than just gas compression as used in the previous DOE/NETL study. The carbon capture from the syngas for this Calibration Case at 88.2% is limited by the CH_4 content of the syngas.

Phillips 66 can design their gasifier system to produce less CH_4 but at the expense of lower gasifier efficiency. An additional case with the cold gas cleanup technology was developed with lower CH_4 content in the syngas such that 90% carbon capture may be achieved. The CH_4

content required in the syngas to reach this higher level of carbon capture is 2.86 mole % versus 4.3 mole % (all on a dry basis) for the previous Calibration Case. The performance for this lower CH_4 content case is also summarized in Table 5. As expected, the overall thermal efficiency dropped, from 32.13% to 31.6% which corresponds to an increase of 1.7% in the heat rate.

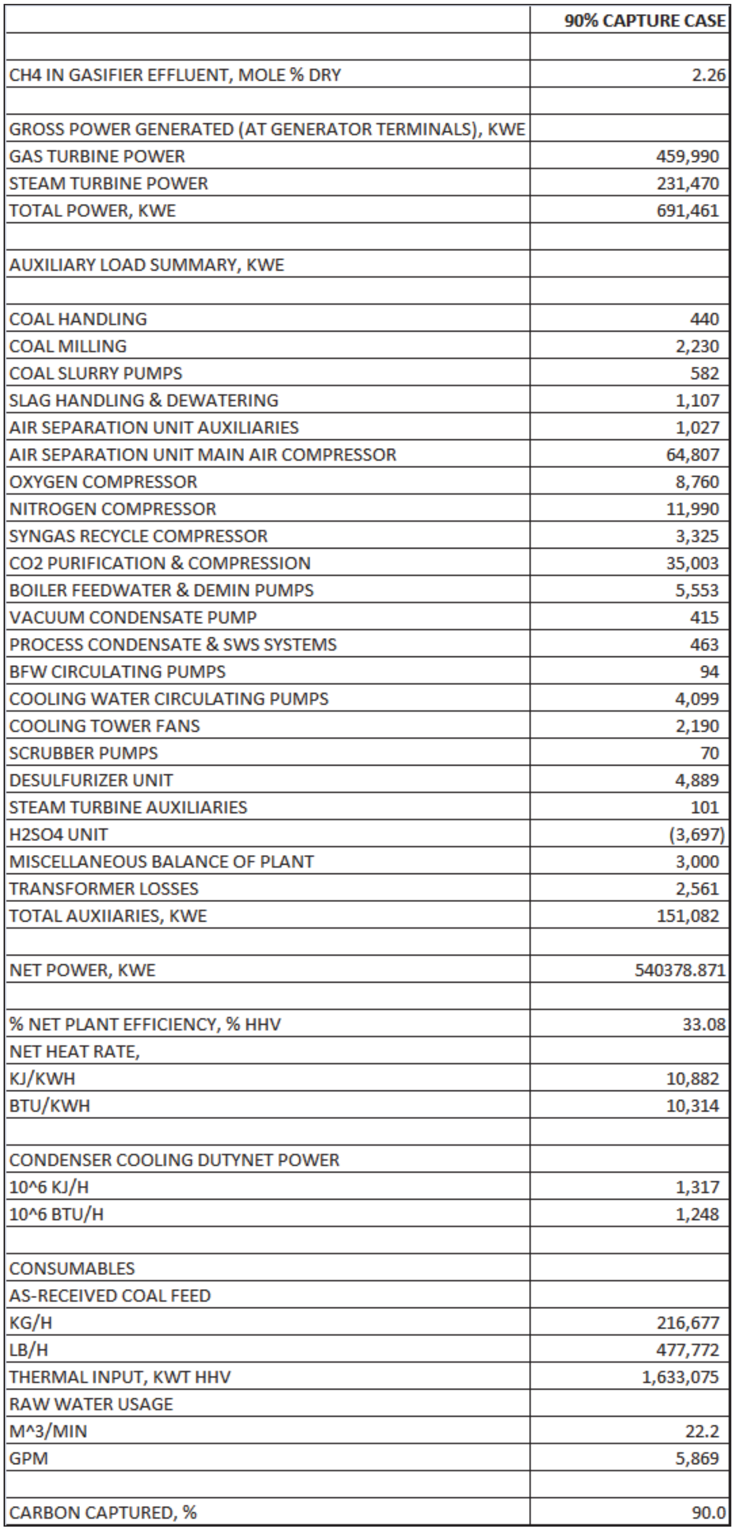
Table 5. Overall Plant Performance Summary Cold Gas Cleanup Cases.



	CALIBRATION CASE	90% CAPTURE CASE
CH4 IN GASIFIER EFFLUENT, MOLE % DRY	4.3	2.86
GROSS POWER GENERATED (AT GENERATOR TERMINALS), KWE		
GAS TURBINE POWER	464,336	461,986
STEAM TURBINE POWER	232,434	229,638
TOTAL POWER, KWE	696,770	691,624
AUXILIARY LOAD SUMMARY, KWE		
COAL HANDLING	440	440
COAL MILLING	2,230	2,230
COAL SLURRY PUMPS	582	582
SLAG HANDLING & DEWATERING	1,107	1,107
AIR SEPARATION UNIT AUXILIARIES	996	1,021
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	62,838	64,427
OXYGEN COMPRESSOR	8,502	8,720
NITROGEN COMPRESSOR	36,378	37,296
SYNGAS RECYCLE COMPRESSOR	3,204	3,296
TAIL GAS RECYCLE COMPRESSOR	1,055	1,074
CO2 COMPRESSOR	19,464	19,888
BOILER FEEDWATER & DEMIN PUMPS	5,173	5,149
VACUUM CONDENSATE PUMP	329	323
PROCESS CONDENSATE & SWS SYSTEMS	366	382
HUMIDIFIER & BFW CIRCULATING PUMPS	108	61
COOLING WATER CIRCULATING PUMPS	4,186	4,172
COOLING TOWER FANS	2,236	2,229
SCRUBBER PUMPS	70	70
SELEXOL UNIT	14,827	15,153
GAS TURBINE AUXILIARIES	1,001	996
STEAM TURBINE AUXILIARIES	101	100
CLAUS & TAIL GAS TREATING AUXILIARIES	199	199
MISCELLANEOUS BALANCE OF PLANT	3,000	3,000
TRANSFORMER LOSSES	2,581	2,562
TOTAL AUXIIARIES, KWE	171,998	175,498
NET POWER, KWE	524,772	516,126
% NET PLANT EFFICIENCY, % HHV	32.13	31.60
NET HEAT RATE,		
KJ/KWH	11,206	11,394
BTU/KWH	10,621	10,799
CONDENSER COOLING DUTYNET POWER		
10^6 KJ/H	1,269	1,250
10^6 BTU/H	1,202	1,185
CONSUMABLES		
AS-RECEIVED COAL FEED		
KG/H	216,677	216,677
LB/H	477,772	477,772
THERMAL INPUT, KWT HHV	1,633,075	1,633,075
RAW WATER USAGE		
M^3/MIN	19.9	19.5
GPM	5249	5159
CARBON CAPTURED, %	88.2	90.0

The performance of the Warm Gas Cleanup Case (with TDA's CO_2 capture system) is summarized in Table 6 (which was developed by further modifying the Aspen Plus® IGCC system model developed for the above Cold Gas cleanup Cases).

Table 6. Overall Plant Performance Summary Warm Gas Cleanup Cases.



	90% CAPTURE CASE
CH4 IN GASIFIER EFFLUENT, MOLE % DRY	2.26
GROSS POWER GENERATED (AT GENERATOR TERMINALS), KWE	
GAS TURBINE POWER	459,990
STEAM TURBINE POWER	231,470
TOTAL POWER, KWE	691,461
AUXILIARY LOAD SUMMARY, KWE	
COAL HANDLING	440
COAL MILLING	2,230
COAL SLURRY PUMPS	582
SLAG HANDLING & DEWATERING	1,107
AIR SEPARATION UNIT AUXILIARIES	1,027
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	64,807
OXYGEN COMPRESSOR	8,760
NITROGEN COMPRESSOR	11,990
SYNGAS RECYCLE COMPRESSOR	3,325
CO2 PURIFICATION & COMPRESSION	35,003
BOILER FEEDWATER & DEMIN PUMPS	5,553
VACUUM CONDENSATE PUMP	415
PROCESS CONDENSATE & SWS SYSTEMS	463
BFW CIRCULATING PUMPS	94
COOLING WATER CIRCULATING PUMPS	4,099
COOLING TOWER FANS	2,190
SCRUBBER PUMPS	70
DESULFURIZER UNIT	4,889
STEAM TURBINE AUXILIARIES	101
H2SO4 UNIT	(3,697)
MISCELLANEOUS BALANCE OF PLANT	3,000
TRANSFORMER LOSSES	2,561
TOTAL AUXIIARIES, KWE	151,082
NET POWER, KWE	540378.871
% NET PLANT EFFICIENCY, % HHV	33.08
NET HEAT RATE,	
KJ/KWH	10,882
BTU/KWH	10,314
CONDENSER COOLING DUTYNET POWER	
10^6 KJ/H	1,317
10^6 BTU/H	1,248
CONSUMABLES	
AS-RECEIVED COAL FEED	
KG/H	216,677
LB/H	477,772
THERMAL INPUT, KWT HHV	1,633,075
RAW WATER USAGE	
M^3/MIN	22.2
GPM	5,869
CARBON CAPTURED, %	90.0

In order to achieve the 90% carbon capture in this Warm Gas Cleanup Case, the CH_4 content in the gasifier effluent had to be further reduced to 2.26 mole % on a dry basis. The resulting efficiency for this case is 33.08% versus 31.6% (both on a coal HHV basis) for the Cold Gas Case when comparing on a consistent carbon capture basis, or a decrease of as much as 4.5% in the heat rate. The plant water consumption goes up, however, by almost 9% on a net kW generated basis for the Warm gas Cleanup Case since a significantly higher amount of water vapor enters the gas turbine along with the syngas and is lost to the atmosphere.

The above results indicate that warm gas cleanup technology coupled with the CO_2 adsorption process being developed by TDA can make a substantial improvement to the thermal performance of an IGCC plant that is designed for near zero emissions, i.e., that includes CO_2 capture.

The system analysis was later updated in the second budget period with the latest results i.e., the final optimized system design for the TDA's CO_2 capture and purification system and a detailed economic analysis (Task 8) was carried out to assess TDA's technology in comparison with competing technologies.

4.4 Task 4. Scale-up of Sorbent Production

In this task, we scaled up the sorbent production using scalable, high throughput manufacturing equipment. We used the BET surface area and room temperature CO_2 isotherm measurements to benchmark the performance of the sorbent formulations prepared in large production batches against the sorbents made at smaller scale.

4.4.1 Scale-up Production

The process for making the sorbent consists of two thermal steps (Figure 28). In the first step the raw materials are heated in trays at 220°C in air to convert them into a char. The more critical step is the carbonization of the char into the product sorbent. In this step, the char is heated under a flow of nitrogen up to 900°C. Our initial samples for this project were made in ceramic trays using a 3" diameter Lindberg tube furnace and our sample sizes were limited to less than 50 g.

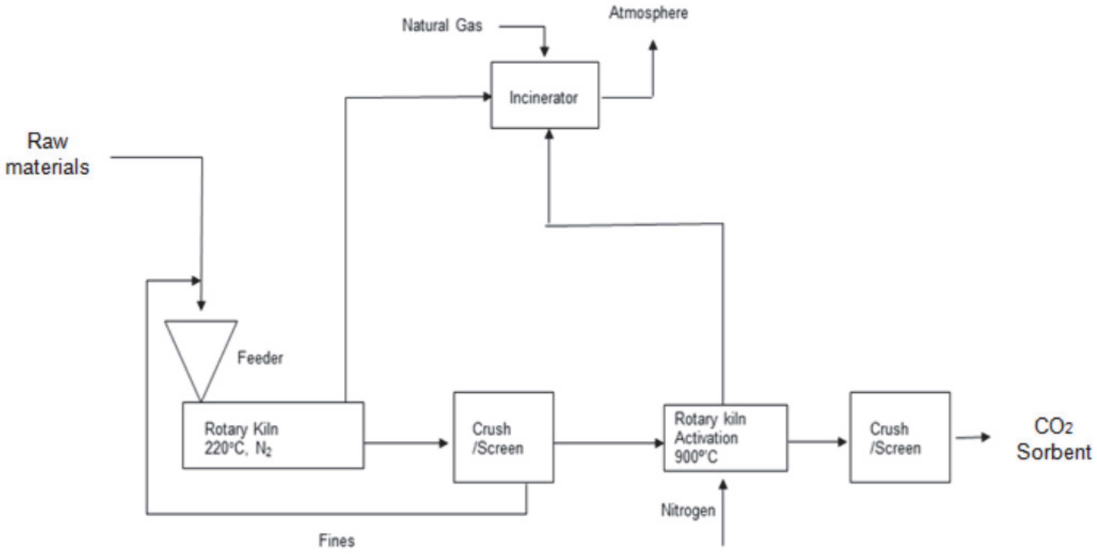


Figure 28. Process for making TDA's CO_2 sorbent.

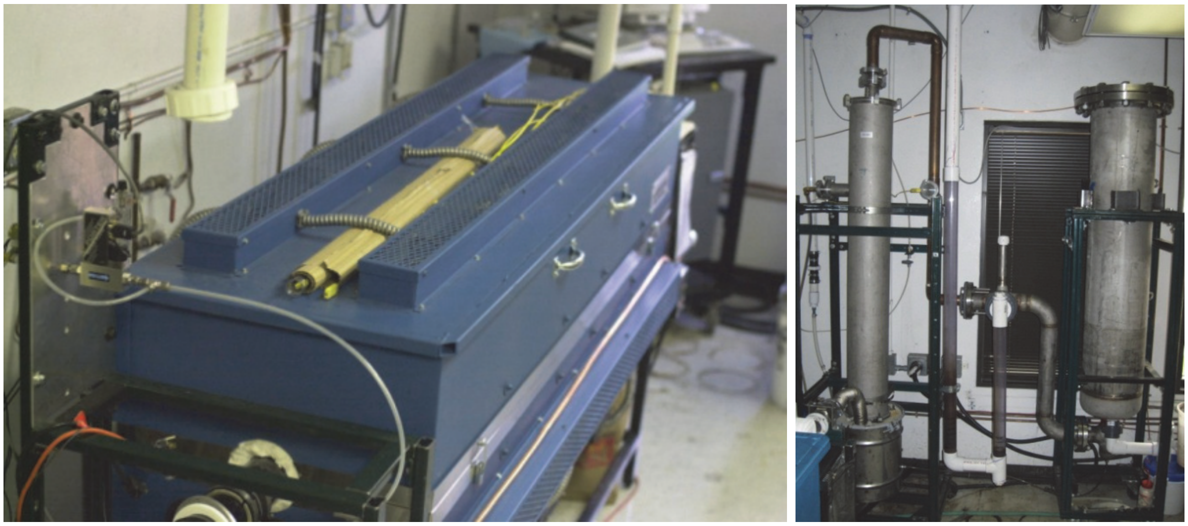
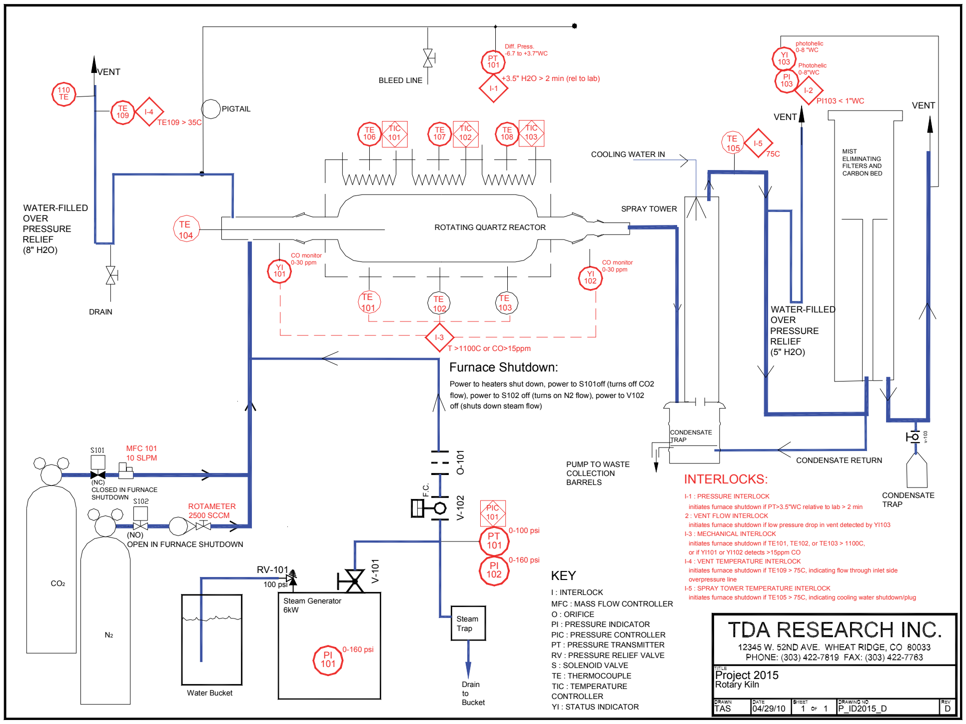


Figure 29. TDA's 11" quartz rotary kiln and the wet scrubber (left) and carbon sorbent bed (right) on TDA's 11" quartz rotary batch kiln.

As part of this task, we scaled up the sorbent production batch size to multi-kilogram quantities, using our 11" rotary kiln with a quartz tube (Figure 29 left). The exhaust of the kiln is plumbed into a wet scrubber followed with a carbon bed as shown in Figure 29 (right). The PID in Figure 30 shows the overall schematic and safety features on the kiln. The exit end of the quartz tube was connected to a wet scrubber with a reservoir to collect the condensable materials. The exhaust then goes through mist eliminators and an activated carbon bed to eliminate smoke and odor before the gases exit the building through the exhaust vent. To overcome the pressure drop due to the carbon bed, a high velocity blower is used in the vent line to reduce the pressure to -3" H_2O. The capacity for our batch kiln is 12 kg of char to give 5 kg of product.



Furnace Shutdown:

Power to heaters shut down, power to S101off (turns off CO2 flow), power to S102 off (turns on N2 flow), power to V102 off (shuts down steam flow)

INTERLOCKS:

I-1 : PRESSURE INTERLOCK
initiates furnace shutdown if PT>3.5"WC relative to lab > 2 min
2 : VENT FLOW INTERLOCK
initiates furnace shutdown if low pressure drop in vent detected by YI103
I-3 : MECHANICAL INTERLOCK
initiates furnace shutdown if TE101, TE102, or TE103 > 1100C, or if YI101 or YI102 detects >15ppm CO
I-4 : VENT TEMPERATURE INTERLOCK
initiates furnace shutdown if TE109 > 75C, indicating flow through inlet side overpressure line
I-5 : SPRAY TOWER TEMPERATURE INTERLOCK
initiates furnace shutdown if TE105 > 75C, indicating cooling water shutdown/plug

KEY

I : INTERLOCK
MFC : MASS FLOW CONTROLLER
O : ORIFICE
PI : PRESSURE INDICATOR
PIC : PRESSURE CONTROLLER
PT : PRESSURE TRANSMITTER
RV : PRESSURE RELIEF VALVE
S : SOLENOID VALVE
TE : THERMOCOUPLE
TIC : TEMPERATURE CONTROLLER
YI : STATUS INDICATOR

TDA RESEARCH INC.
12345 W. 52ND AVE. WHEAT RIDGE, CO 80033
PHONE: (303) 422-7819 FAX: (303) 422-7763

TITLE				
Project 2015 Rotary Kiln				
DRAWN	DATE	SHEET	DRAWING NO.	REV
TAS	04/29/10	1 OF 1	P_ID2015_D	D

Figure 30. PID for 11" quartz rotary batch kiln.

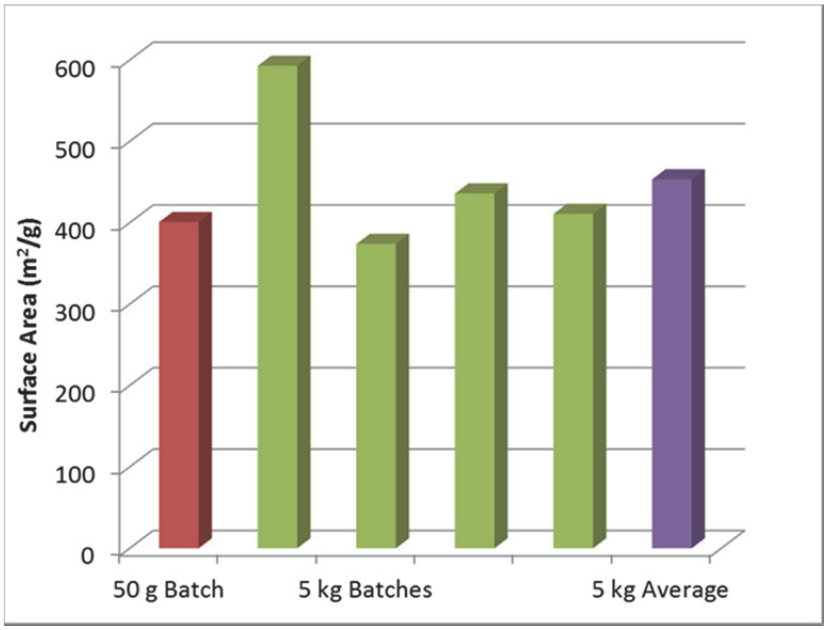


Figure 31. BET surface area comparison of large batch preparations against the small batch.

4.4.2 BET Surface Area Measurements

After preparing the sorbents in large batches, we measured their BET surface area in our Micromeritics Gemini surface area analyzer (Figure 5). Figure 31 shows the BET surface area for AMS-19 prepared in the large rotary at 5 kg batch size against the average from the small batch preparations. In these tests we were optimizing the preparation procedure to achieve a tighter control on the surface area of the sample. The later 5 kg batches had a surface area closer to the 50 g batch, as shown in Figure 31.

4.4.3 Low Temperature Measurements

TDA has an automated constant volume apparatus (CVA) for adsorption isotherm measurements from vacuum to pressures up to 1.8 atm. We used this apparatus to measure the adsorption isotherms at room temperature for characterizing the CO_2 adsorption capacity of the large production batches. The CVA allows us to measure single component gas-solid equilibrium and kinetics (adsorption characteristics). The constant volume adsorption process is a static measurement technique where the pressure change due to adsorption in a closed constant volume system is measured with a pressure transducer. The adsorbed amount is calculated by making a material balance using any

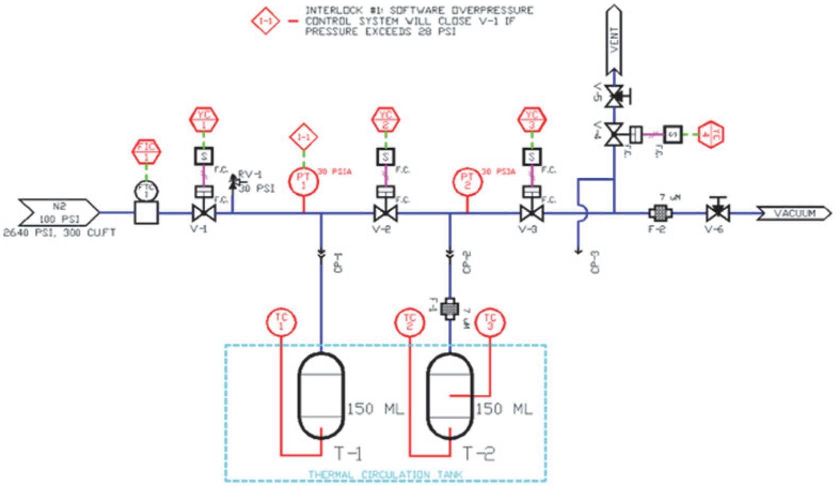


Figure 32. P&ID of TDA's constant volume apparatus.

standard thermodynamic equation of state such as the "ideal gas law". Figure 32 shows the P&ID and Figure 33 is a picture of the constant volume apparatus. The set up consists of two cylindrical chambers or cells (dose and sample (test) cells) of known volume connected by a pneumatic ON/OFF valve. A few grams of the adsorbent (about 2-5 g) are placed in the sample cell. The feed gas from the gas cylinder is then filled into the dose cell through the mass flow controller to a pre-set pressure at which point the gas inlet valve (#V-1) is closed. Then the ON/OFF valve (#V-2) between the two cells is opened to provide a step change in the pressure of the sample cell. The size of the step depends on the time of opening of the valve and the initial difference in pressure (∆P) between the two sides. This valve can be left open until equilibration of the pressures between the two sides to operate in pressure equalization mode. The pressure change after opening the valve is monitored by the data logging computer to generate the adsorption equilibrium and kinetics data.

Figure 33. Picture of TDA's constant volume apparatus.

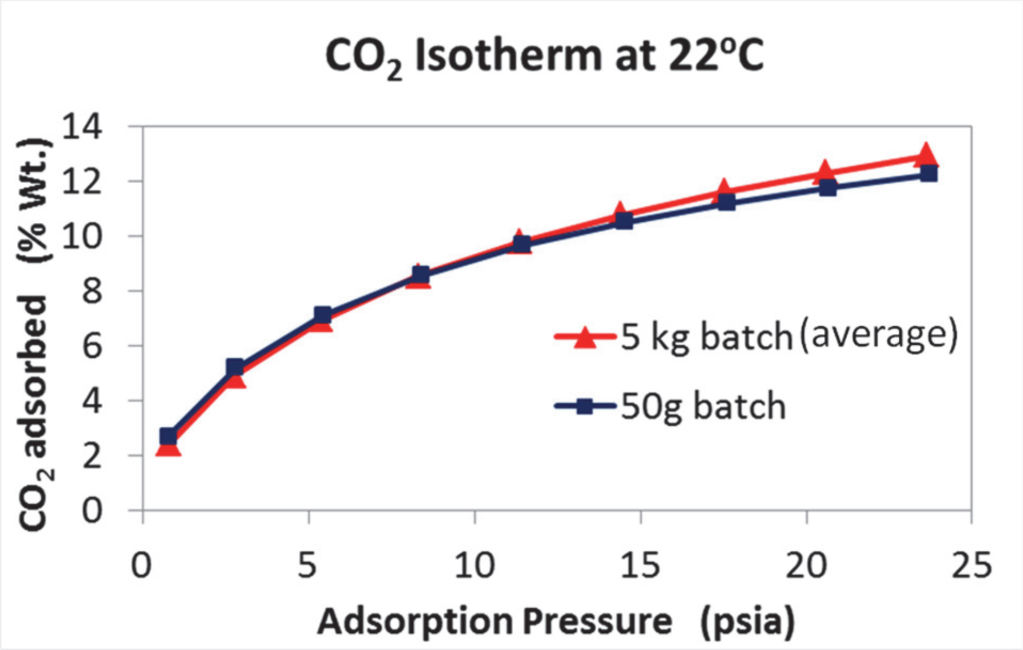


Figure 34. CO_2 adsorption isotherm comparison of large batch preparations against the small batch.

We measured CO_2 adsorption isotherms at room temperature (i.e., 22°C). The adsorption kinetics are fast and the CO_2 uptake is rapid, indicating a high diffusion rate of 1.3 x 10^{-3} (D/r_c^2, s^{-1}). The averaged CO_2 adsorption isotherm for the large production batches at room temperature along with the isotherm for the small batch is provided in Figure 34.

4.4.4 Sorbent Cost Analysis

The process diagram is shown in Figure 35, followed by the material balance for this process (Table 7). The raw materials and water are fed onto the Teflon belt of the tunnel kiln as a syrup. The mixture is heated to 220°C to decompose the sugar into a black char. At the end of the kiln the char is removed, crushed and screened. The fines are returned to the feeder and the granular material is activated. Overall yield is 14% (minus water).

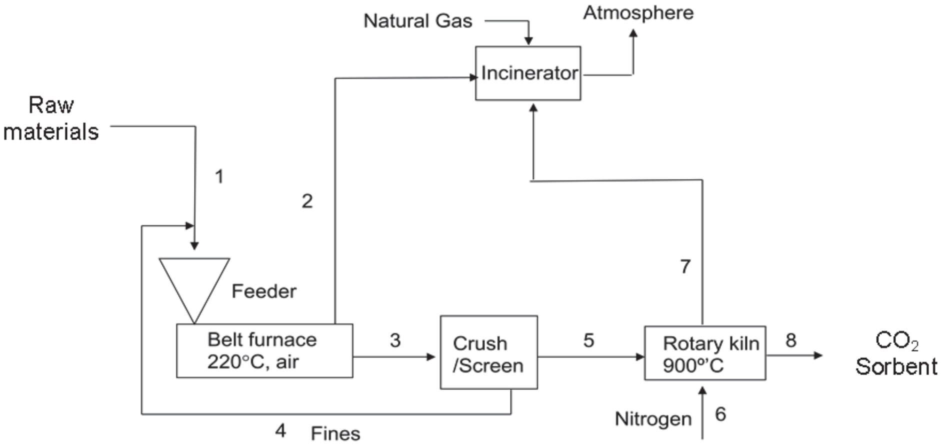


Figure 35. Process for producing TDA's CO_2 sorbent in a continuous rotary kiln.

Table 7. Materials balance for the various steams to make 1 kg of sorbent (AMS-19) using TDA's scaled-up sorbent production process shown in Figure 35.

Component (kg)	1	2	3	4	5	6	7	8
Raw material 1	5.0							
Raw material 2	0.8							
Raw material 3	1.2							
Water	3.6							
Char			3.8					
Fines	0.6			0.6				
Crushed char					3.2			
CO_2 Sorbent								1
Volatiles		6.8					2.2	
Nitrogen						3.0	3.0	
Total kg	11.2	6.8	3.8	0.6	3.2	3.0	5.2	1

The lowest cost that is possible for the sorbent would be based solely on the starting materials cost. The total cost of the raw materials is calculated to be \$3.40 per kg. The product costs are based on an overall yield of 14% yield. Based on prior work that MWV has done for us to estimate the cost of our ultra-capacitor carbons, we carried out the economic analysis on the process based on an annual sorbent production capacity of 5000 tons/year (Table 8). Currently the production cost is estimated to be \$5.30/kg (\$2.41 per lb) based on current raw materials costs, which accounts for 64% of the total production costs.

Table 8. Estimated production costs for TDA's CO_2 Sorbent.

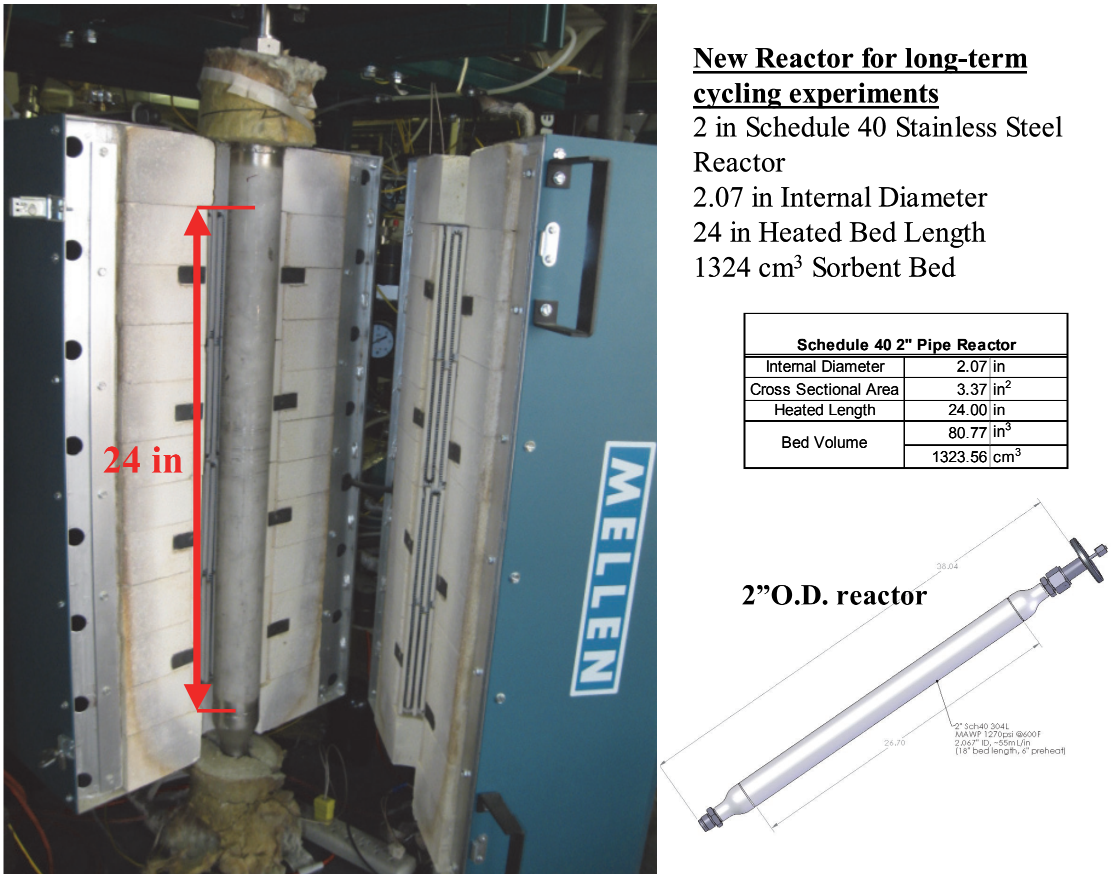
Basis 5,000 tons/year	Annual Production Costs ($/kg)
Raw Materials	3.40
Labor	0.24
Utilities, etc.	1.26
Capital Recovery, Manufacturing Indirects and Profit	0.40
Total	5.30

4.5 Task 5. Long-term Sorbent Cycling Experiments

In this task, we carried out long-term cycling experiments under simulated synthesis gas to assess the impact on sorbent's CO_2 capacity and removal efficiencies. The sorbent maintained its capacity and removal efficiencies over 11,650 cycles. We also carried out tests in the presence of contaminants such as H_2S and the sorbent maintained its capacity in the presence of H_2S.

4.5.1 Experimental Set-up

We modified an existing sorbent testing apparatus with fixed-bed reactor to carry out long-term cycling experiments with the sorbent under representative synthesis gas conditions. We



New Reactor for long-term cycling experiments

2 in Schedule 40 Stainless Steel Reactor

2.07 in Internal Diameter

24 in Heated Bed Length

1324 cm^3 Sorbent Bed

Schedule 40 2" Pipe Reactor		
Internal Diameter	2.07	in
Cross Sectional Area	3.37	in^2
Heated Length	24.00	in
Bed Volume	80.77	in^3
	1323.56	cm^3

2"O.D. reactor

Figure 36. Long-term cycling apparatus with the large 2"O.D. reactor.

fabricated a larger reactor that can hold up to 1.3 L of sorbent, which allowed us to reduce the dead volume to sorbent volume ratio so as to minimize the impact of the dead spaces on the breakthrough profile. The reactor is made of a 2" schedule 40 stainless steel tube with end caps and fittings welded on either ends. Figure 36 shows the picture of the long-term cycling apparatus with the large 2" O.D. reactor. Based on the information provided by Phillips 66, we selected a representative gas composition (using the equilibrium composition downstream of a water-gas-shift reactor to convert as much carbon as possible to CO_2). Table 9 shows the synthesis gas composition used in the long-term cycling tests. We carried out regenerations at a lower pressure to provide pressure swing and used mixtures of H_2 and H_2O to simulate the steam purge used in the actual system since the CO_2 analyzer cannot handle full steam. However, in these tests we maintained the H_2O partial pressure to be same as the steam purge pressure in the actual system.

Table 9. Simulated synthesis gas composition and steam purge used in long-term cycling tests.

	Synthesis Gas	Steam Purge
Temperature	240°C	240°C
Pressure	500 psig	300 psig
Composition		
H_2	15.0%	50.0%
CO_2	48.1%*	
H_2O	36.3%	50.0%+
CO	0.6%	

* adjusted for gasifier operation at 750 psia

\+ adjusted for purge with 100% steam at 150 psia

4.5.2 Long-term Cycling Results

In the long-term cycling experiments we carried a simple pressure swing adsorption cycle with 4-steps: pressurization, adsorption, de-pressurization and purge/desorption. Typical test data from an adsorption/ regeneration cycle is shown in Figure 37. During the pressurization step, the synthesis gas is used to raise the bed pressure to the adsorption pressure (please note there is no flow through the analyzer during this portion of the cycle) and after the bed reaches the adsorption pressure (500 psig) the CO_2 free synthesis gas is produced as adsorption product indicating complete removal of CO_2. Next, as the CO_2 starts to breaks through from the bed the pressure is reduced to

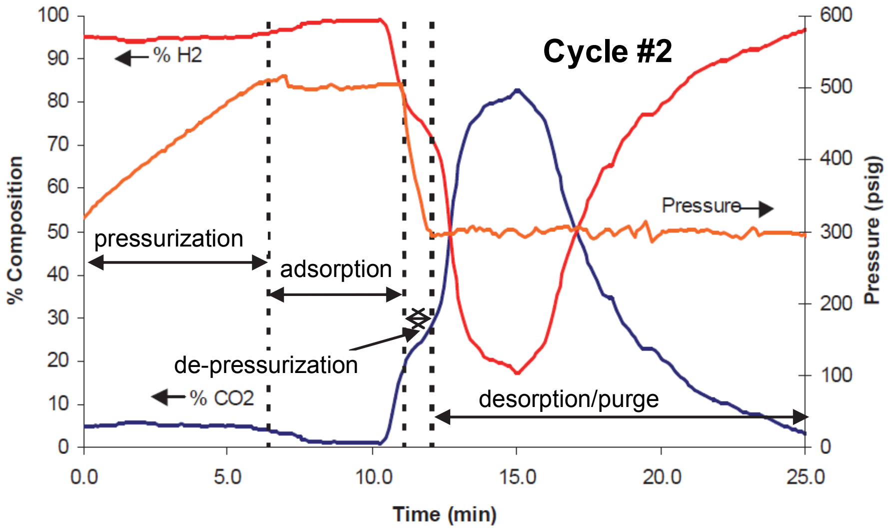


Figure 37. Data from a typical cycle from the long-term cycling experiments carried out at 240°C.

the desorption pressure (300 psig). Steam at 150 psig is introduced as a purge gas and the adsorbed CO_2 is released from the bed producing concentrated CO_2 stream that can be further purified if needed. The CO_2 working capacity of the sorbent is calculated the amount of CO_2 captured by the sorbent during a cycling per kg of sorbent and the CO_2 removal rate is calculated as the fraction of CO_2 captured.

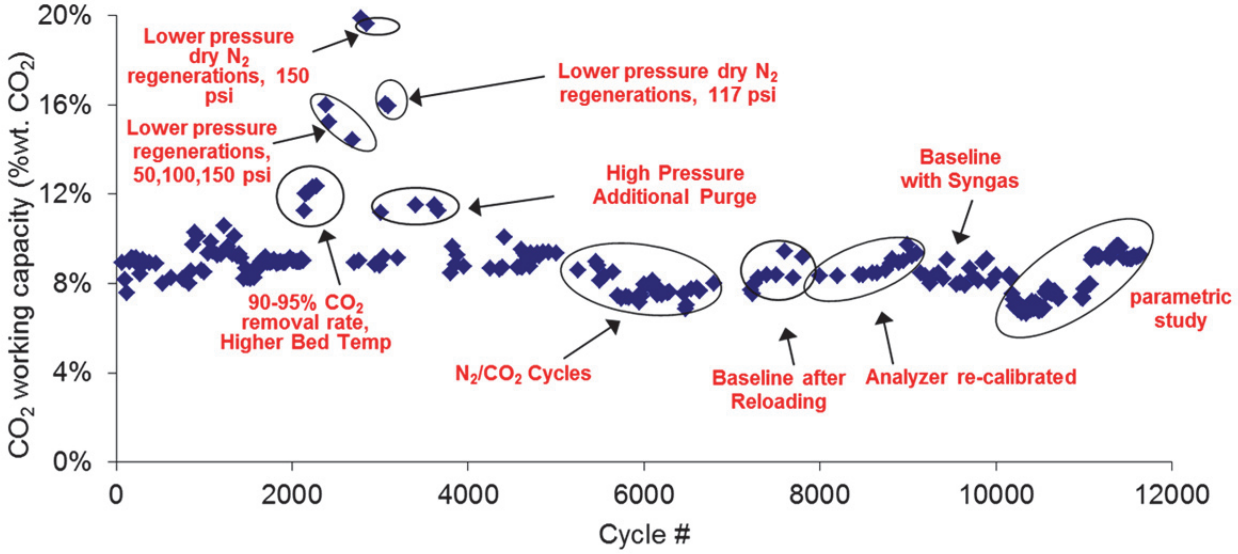


Figure 38. TDA's CO_2 sorbent's working capacity under cycling in simulated synthesis gas.

We carried out more than 11,650 cycles and the sorbent showed stable performance (Figure 38). During these tests, in some experiments, we varied the adsorption and desorption (regeneration) conditions by changing the temperature, pressure and purge gas and volume.

In cycles 2100-2163, we increased the adsorption bed temperature to 260°C. This increased the working capacity of our sorbent from about 9% wt. at the baseline conditions up to 12.4% wt. CO_2. Between cycles 2164-2310, we decreased the regeneration pressure to 50 psi, 100 psi, and 150 psi. These regeneration pressures all resulted in a capacity of about 15-16% wt. CO_2. In cycles 2672-2716, we ran normal adsorption cycles with dry nitrogen regenerations at 150 psi. This resulted in the highest capacities, about 20% wt. CO_2. In cycles 2977-3014, we ran normal adsorption cycles with dry nitrogen regenerations at 117 psi. This resulted in higher CO_2 capacity of about 16% wt. CO_2. In cycles 3050-3096, we ran baseline conditions but added a 5 minute dry nitrogen purge at 500 psi. This resulted in capacities of about 11.5% wt. CO_2. We ran $N_2/CO_2/H_2O$ from cycles 3614-4425, which resulted in very similar capacities to baseline conditions (8-10 wt. CO_2.)

In the cycles between 5,500 to 7,000 cycles the apparent CO_2 capacities and removal efficiency were lower due to drift in the analyzer. Once we re-calibrated the analyzer using calibration gas mixtures the measured sorbent performance was back to original capacity of about 9% wt. CO_2 at our standard test conditions and the sorbent maintained its capacity over 10,000 cycles. After completing the 10,000 cycles we explored some more parametric study to analyze the impact of higher adsorption temperatures (280°C). Finally, we went back to standard test condition and the sorbent achieved a capacity of 9.% wt. CO_2 between 11,000 and 11,650 cycles. The CO_2 removal rate was stable over 10,000 cycles achieving over 95% removal efficiency as shown in Figure 39. The CO_2 breakthrough during the 10,000 cycles is shown in Figure 40.

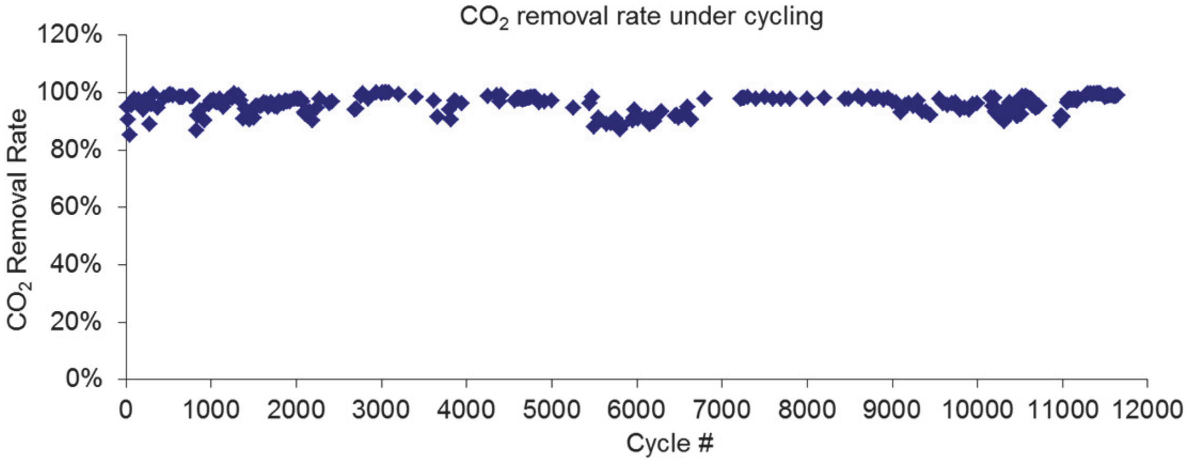


Figure 39. TDA's CO_2 sorbent's CO_2 removal rate under cycling in simulated synthesis gas.

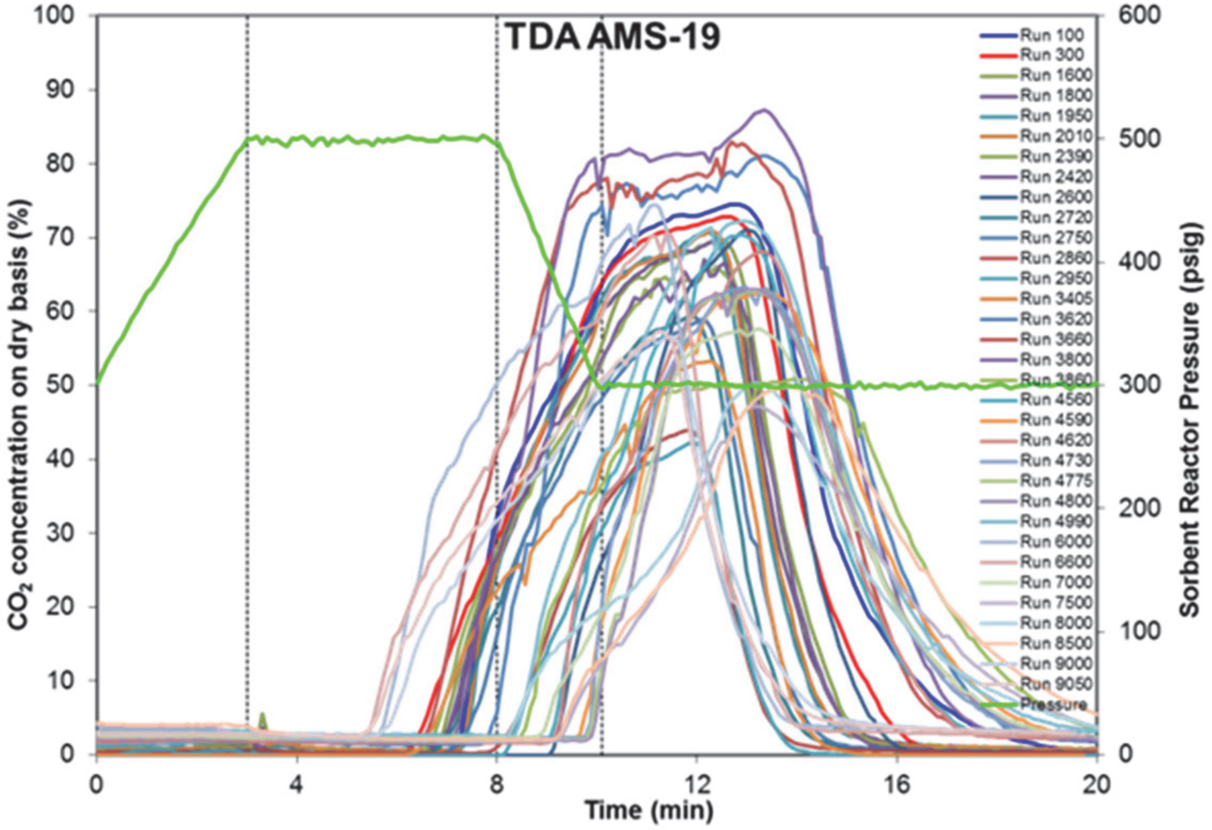


Figure 40. CO_2 breakthrough profile from the long-term cycling experiments.

4.5.3 Impact of H_2S

We carried out multiple cycle fixed bed adsorption tests with TDA's CO_2 sorbent (AMS-19) using simulated synthesis gas stream containing 43.4% vol. CO_2, 15.1% vol. H_2, 36.5% vol. H_2O at 240°C and 500 psig for adsorption and a mixture of 50% vol. H_2O in H_2 for desorption/purge. These tests were carried out in a different experimental setup and we carried out concentration swing cycles instead of the pressure swing cycles. The sorbent maintained its CO_2 capacity when we introduced 10 ppmv H_2S achieving similar capacity in the presence of H_2S. The impact of 10 ppmv H_2S on the sorbent's CO_2 capacity is shown in Figure 41.

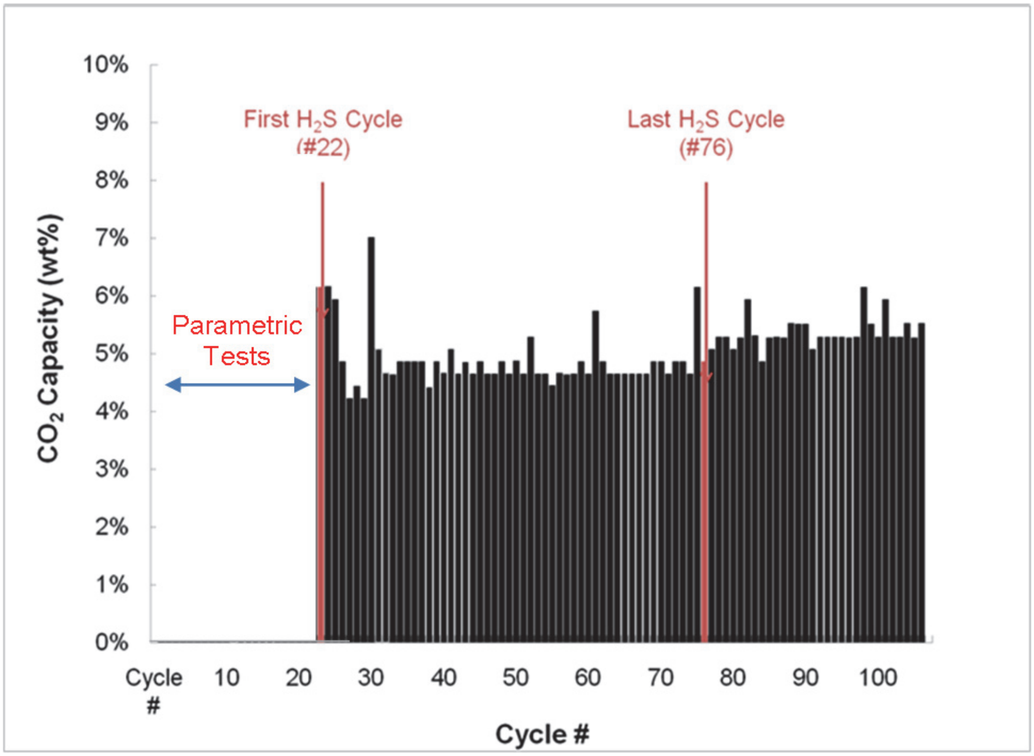


Figure 41. CO_2 working capacity in the presence of 10 ppmv H_2S at 240°C and 500 psig under simulated synthesis gas

We also carried out tests in the presence of higher levels of H_2S up to 300 ppmv H_2S. The sorbent maintained a stable performance in the presence of 300 ppmv H_2S. We observed similar breakthrough for CO_2 in the presence and absence of 300 ppmv H_2S (Figure 42 left). The sorbent maintained its CO_2 capacity when we introduced 10 ppmv H_2S achieving similar capacity in the presence of H_2S. The impact of 300 ppmv H_2S on the sorbent's CO_2 capacity is shown in Figure 42 (right).

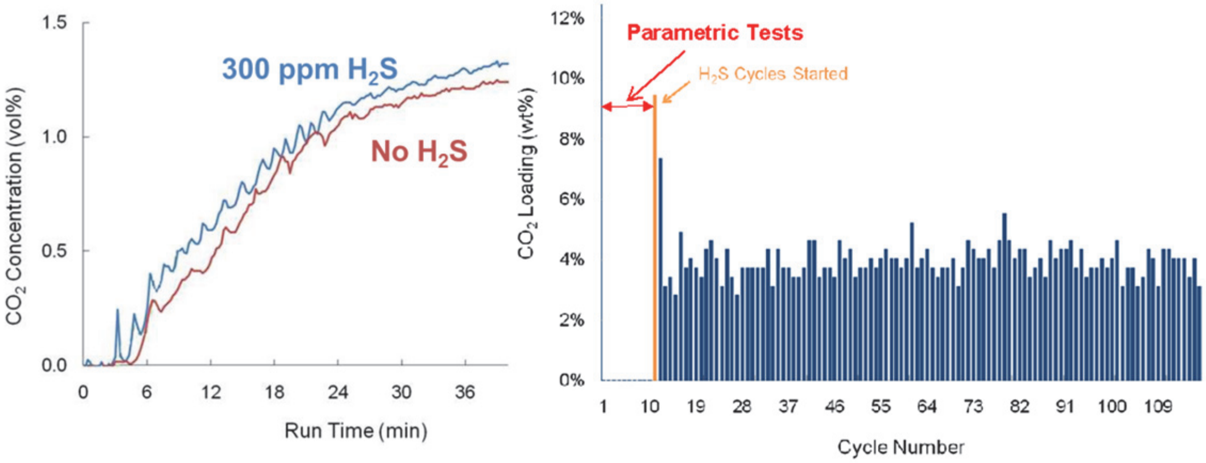


Figure 42. CO_2 breakthrough and working capacity in the presence of 300 ppmv H_2S.

4.6 Task 6. Design of PSA System

In the original proposal, we proposed a moving-bed reactor system for CO_2 removal using our sorbent. Based on the Year 1 results, after careful investigation of the CO_2 adsorption/desorption characteristics of the sorbent (i.e., the sorbent achieved higher working capacities under a combination of pressure and concentration swing regeneration), we decided to house the sorbents in fixed bed reactors, applying a combination of pressure and concentration swing (i.e. steam purge) to regenerate the sorbent (higher working capacities allowed us to use the fixed bed reactors and eliminated the need for more complex moving beds with lock hoppers to adjust for the pressure difference between adsorption and regeneration steps). Hence, in this task (Task 6: Design of Moving Bed Reactors), we carried out the computer simulations of the PSA cycle scheme to optimize the operating parameters and design the reactor system. The simulations allowed us to optimize the number of reactors and the cycle sequence that we will need to achieve the targeted 90% carbon capture with minimum power and steam consumption for the sorbent regeneration.

4.6.1 PSA Cycle Model and Simulations

Figure 43 shows a typical PSA cycle scheme with three pressure equalizations for pre-combustion CO_2 capture system using TDA's sorbent. The computer simulation program uses mathematical models to represent each of the steps shown in Figure 43. This model assumes axial dispersed plug flow of a binary gas mixture through a packed adsorbent bed with spherical adsorbent particles (H_2, CO_2). We have used the parameters obtained from bench-scale fixed bed experiments for CO_2 adsorption (from Task 2) to carry out the binary simulations. The sorbent bed is assumed to be non-isothermal and the external film diffusional resistance is assumed to be negligible compared to the resistance in the pores of the sorbent. The axial pressure drop is neglected and gases are assumed to be ideal.

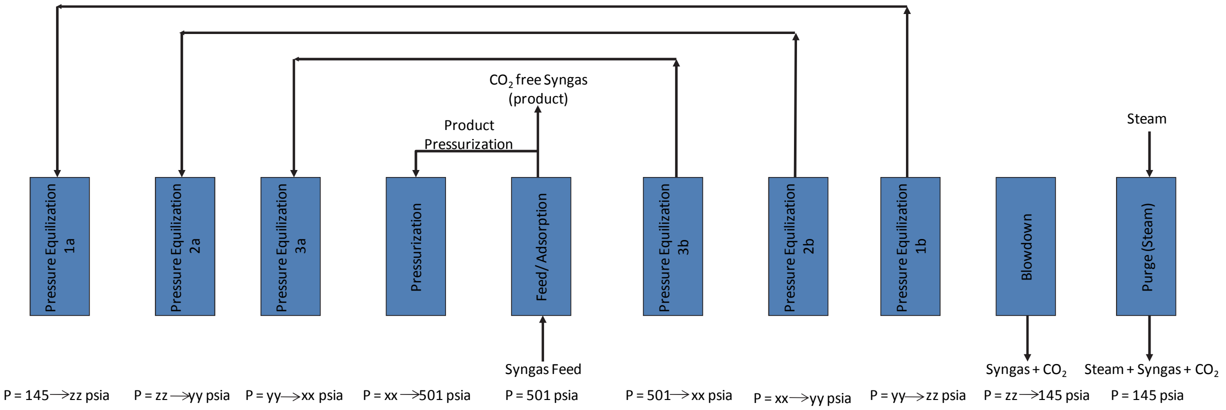


Figure 43. Typical cycle scheme for pre-combustion CO_2 capture that includes three pressure equalization steps to improve the H_2 and CO_2 recoveries.

The equilibrium relations for both the components are represented by extended binary Langmuir-Freundlich isotherms. The average diffusional time constants are estimated from the bench-scale kinetic data from the isotherm unit. The exponential pressure profile (history) is assumed for the variable pressure steps with the time constant chosen according to the thumb rule described by Farooq et al. (1993).

For the initial simulations we used a more simplified cycle scheme to validate the model. The simplified cycle scheme is shown in Figure 44. The simplified model equations for isothermal conditions are reported below more details about the model and computer simulations can be found in the Ph.D. dissertation by Jayaraman (2004).

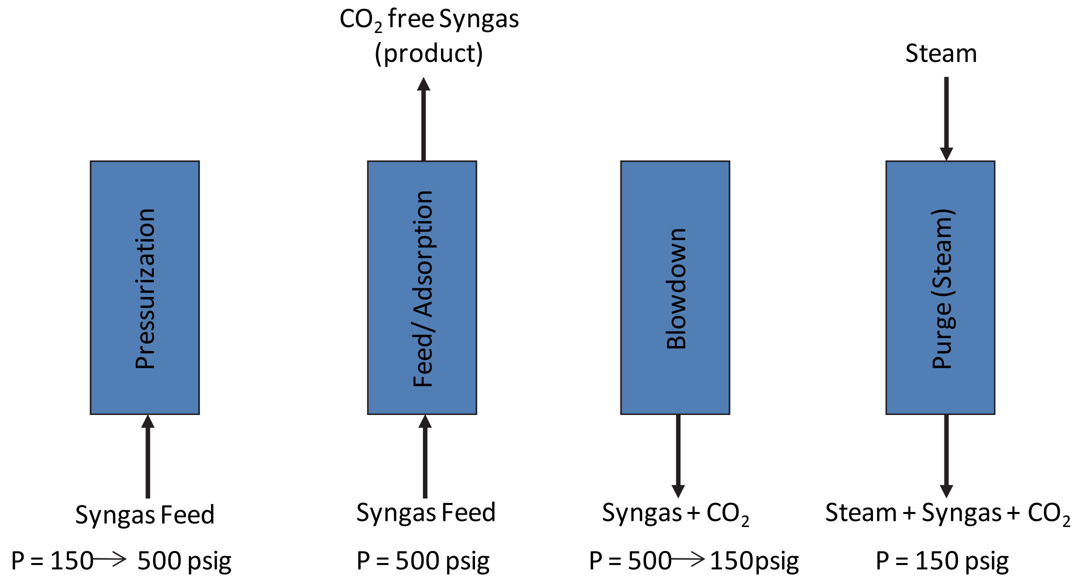


Figure 44. Simplified PSA cycle scheme without pressure equalizations.

The fluid phase mass balance for component *k* is given below:

$$\varepsilon_t \frac{\partial y_k}{\partial t} - \varepsilon D_{ax} \frac{\partial^2 y_k}{\partial z^2} + \varepsilon \frac{\partial (u y_k)}{\partial z} + \frac{\rho_b RT}{P} \frac{\partial \overline{q_k}}{\partial t} + \frac{\varepsilon_t y_k}{P} \frac{dP}{dt} = 0 \qquad (9)$$

The overall material balance is given by:

$$\varepsilon \frac{\partial u}{\partial z} = -\frac{\rho_b RT}{P} \sum_{k=1}^{2} \frac{\partial \overline{q_k}}{\partial t} - \frac{\varepsilon_t}{P} \frac{dP}{dt} \qquad (10)$$

The particle phase mass balance for component *k* as given by the intraparticle diffusion equation for a sphere:

$$\frac{\partial q_k}{\partial t} = \frac{D_{e,k}}{r^2} \frac{\partial}{\partial r} \left(\frac{1}{r^2} \frac{\partial q_k}{\partial r} \right) \qquad (11)$$

with boundary conditions:

$$\frac{\partial q_k}{\partial r} = 0 \quad at\ r = 0 \qquad (12)$$

$$-D_{e,k}\left.\frac{\partial q_k}{\partial r}\right|_{r=R}=a_k\left(q_k\big|_{r=R}-q_k^*\right)\quad at\ r=R \tag{13}$$

where q_k^* is the equilibrium amount adsorbed at the surface of the crystal and can be calculated using the extended binary Langmuir isotherm:

$$q_k^*=\frac{q_{mk}B_kP_k}{1+\sum_{j=1}^{2}B_jP_j} \tag{14}$$

The volume averaged adsorbed phase concentration $\overline{q_k}$ is given by

$$\overline{q_k}=\frac{3}{R^3}\int_0^R q_k r^2 dr \tag{15}$$

The boundary conditions for the fluid phase mass balance are given below for Cycle (I):

(1) Pressurization Step:
at z = 0, $y_k = y_{f,k}$
at z = L, u = 0
$P = P(t) = P_{DES} + (P_H - P_{DES})(t/\tau_p)$

$$D_{ax}\left.\frac{\partial y_k}{\partial z}\right|_{z=0}=u_H\left(y_k\big|_{z=0}-y_{H,k}\right) \tag{16a}$$

$$\left.\frac{\partial y_k}{\partial z}\right|_{z=L}=0 \tag{16b}$$

(2) High Pressure Feed Step:
at z = 0, $y_k = y_{f,k}$ $u = u_f$
$P = P_H$

$$D_{ax}\left.\frac{\partial y_k}{\partial z}\right|_{z=0}=u_H\left(y_k\big|_{z=0}-y_{H,k}\right) \tag{17a}$$

$$\left.\frac{\partial y_k}{\partial z}\right|_{z=L}=0 \qquad \text{(17b))}$$

(3) Countercurrent Blowdown Step:
at z = L, u = 0
$P = P(t) = P_{CBD} + (P_H - P_{CBD})(t/\tau_{bd})$

$$\left.\frac{\partial y_k}{\partial z}\right|_{z=0} = 0$$

(18a)

$$\left.\frac{\partial y_k}{\partial z}\right|_{z=L} = 0$$

(18b)

(4) Countercurrent Desorption / Purge Step:
at z = L, $y_k = y_{p,k}$
at z = L, $u = u_L$
$P = P(t) = P_{DES} + (P_{CBD} - P_{DES})(t/\tau_{des})$

$$-D_{ax}\left.\frac{\partial y_k}{\partial z}\right|_{z=L} = u_L(y_{L,k} - y_L|_{z=L}); \qquad (u_L<0) \quad (19a)$$

$$\left.\frac{\partial y_k}{\partial z}\right|_{z=0} = 0 \qquad (19b)$$

Table 10. Summary of simulation parameters used.

Sorbent Bed Parameters

Bed Length	0.51	meters
External Porosity	0.4	
Radius of the Adsorber	0.0635	meters
Bed (L/D)	4.0	
Bed Density	354	kg/m^3
Shape of Pellets	spherical	
Pellet Porosity	0.57	
Radius of Pellets	3.0E-04	meters
Number of Grids in the Bed	100	
Axial Dispersion Coefficient	5.38E-06	m^2/sec
Heat Capacity of the Bed	930	J/kg/K
Effective Heat Conductivity	0.11	W/m/K
Heat Transfer Coefficient	5	$W/m^2/K$
Wall Temperature	240	°C

CO_2 Adsorption Parameters

Langmuir Coefficient (q_s)	386.4	mmol/g
Langmuir Coefficient (B)	4.15E-04	1/atm
Langmuir Coefficient (n)	0.869	
Diffusion Coefficient (D/R^2)	1.32E-03	1/s
Reference Temperature for B	240	°C
Heat of Adsorption (ΔH)	4.8	kcal/mol

The above partial differential equations were solved by an implicit finite difference method using a Crank-Nicolson scheme. In the simulation 100 grid points are used in the bed with the convergence criterion set at 1 x 10^{-3}. The FORTRAN PSA simulation code developed by Sun et al. (1996) is used. In most simulations the steady state was reached within 50 cycles, so carried out simulations up to 100 cycles (Figure 45).

Table 10 shows the list of the parameters used in the binary simulations. Figure 46 shows the gas phase H_2 wave fronts as it travels (from left to right) through the sorbent bed during the course of the Feed (adsorption) step. Similar wave fronts are calculated for each of the steps for both H_2 and CO_2. The computer simulations provides the bed profiles at various steps in the PSA cycle and can be used to optimize the step times, step pressures and purge volumes etc.

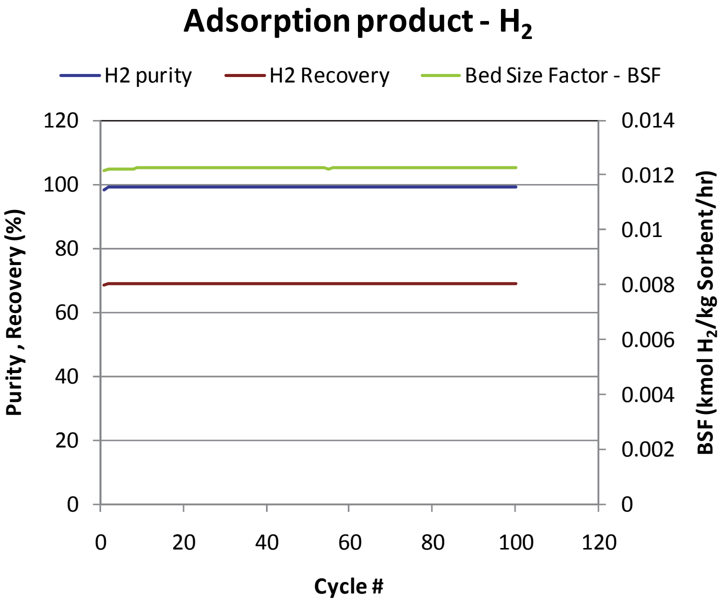


Figure 45. Simulation results showing steady state.

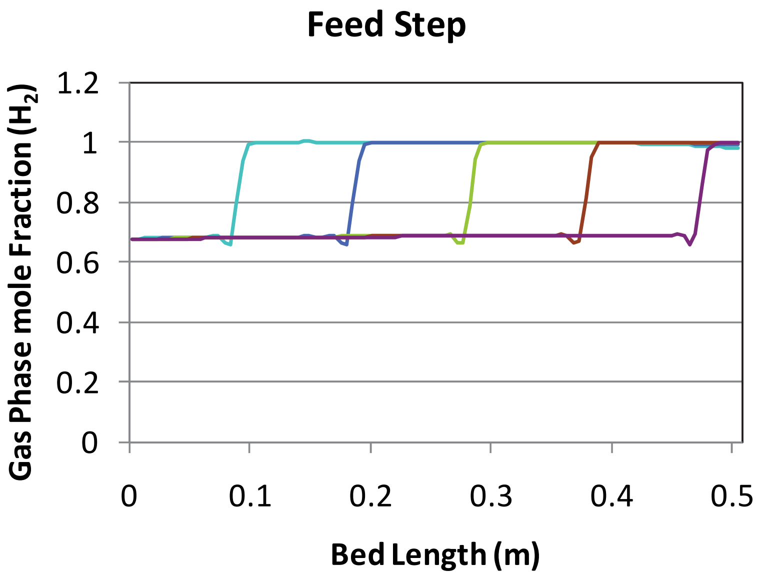


Figure 46. H_2 wave fronts during feed step.

We optimized step times for the simplified cycle scheme and the system performance results for the optimized condition are reported in Table 11.

The CO_2 working capacity calculated by the mathematical model based computer simulations for a simple 4 step cycle is 7.3% wt. CO_2, which is similar to the results obtained in the long-term cycling results. We extended the model to include more complex PSA cycle steps such as pressure equalizations, product pressurization, co-current depressurization. This more complicated cycle sequence will decrease the working capacity of the sorbent. However, the H_2 recovery will increase providing high purity CO_2 during desorption eliminating or simplifying the need for downstream purification of the CO_2 stream before pressurization to 2,200 psig for transport and storage. The recovery and subsequent combustion of H_2 in the gas turbine also increases the overall cycle efficiency.

Table 11. System performance under optimum cycle parameters for simplified scheme (cycle #1).

Two bed (No PE)	
Feed Pressurization - Cycle #1	
H_2 Purity (%)	99.23942
H_2 recovery (%)	69.06249
kmol H_2/kg sorbent/hr	0.012244
CO_2 purity (%)	60.1531
CO2 recovery (%)	98.66
kmol CO_2/kg sorbent/hr	0.00828
Working capacity (% wt. CO_2) - Based on High Pressure H_2 produced	4.985%
Working capacity (% wt. CO_2) - Based on Low Pressure CO_2 produced	7.286%

4.6.2 Optimization of PSA Cycle Scheme

We used the computer simulations to optimize the following PSA cycle parameters:

- Feed Vs Product Pressurization
- Number of Equalization steps
- Steam Purge Volume and Time
- Desorption Pressure

Table 12. Optimized cycle parameters for feed and product pressurizations with product end pressure equalizations.

Feed Pressurization - pdt. end Equalization

	1 PE	2 PE	3 PE
H_2 Purity (%)	99.25	98.25	98.98
H_2 recovery (%)	80.92	85.72	87.48
kmol H_2/kg sorbent/hr	0.01019	0.01016	0.00961
CO_2 purity (%)	70.97	76.13	78.71
CO2 recovery (%)	98.45	96.11	97.69
kmol CO_2/kg sorbent/hr	0.00587	0.00540	0.00509
Working capacity (% wt. CO_2) - Based on Low Pressure CO_2 produced	6.031%	5.541%	5.223%

Product Pressurization - pdt. end Equalization

	1 PE	2 PE	3 PE
H_2 Purity (%)	99.25	99.62	99.43
H_2 recovery (%)	80.49	85.36	87.99
kmol H_2/kg sorbent/hr	0.00999	0.00978	0.00966
CO_2 purity (%)	70.46	76.20	79.52
CO2 recovery (%)	98.46	99.18	98.69
kmol CO_2/kg sorbent/hr	0.00577	0.00537	0.00512
Working capacity (% wt. CO_2) - Based on Low Pressure CO_2 produced	5.929%	5.511%	5.254%

Figure 47 shows the PSA cycle performance data for TDA's CO_2 sorbent with feed and product end pressurizations. We observed that product pressurization provides slightly better performance than feed pressurization when number of pressure equalizations is more than 3. Figure 48 shows the impact of number of equalizations on the synthesis gas recovery (i.e., the percentage of syngas recovered at higher pressure) and the working capacity of the sorbent. We observed that as the number of equalizations increases the working capacity of the sorbent decreases slightly while the synthesis gas recovery increases even more sharply. Increase in synthesis gas recovery directly correlates to a decrease in operating costs and increase in CO_2 (desorption) product purity while a decrease in working capacity correlates to a direct increase in fixed cost (equipment cost). In the CO_2 capture system operating costs are a big component of the total (i.e., \$/ton CO_2 captured). Hence, having 3 pressure equalizations is identified as the optimal in our design as more equalizations will further increase the complexity of the PSA cycle.

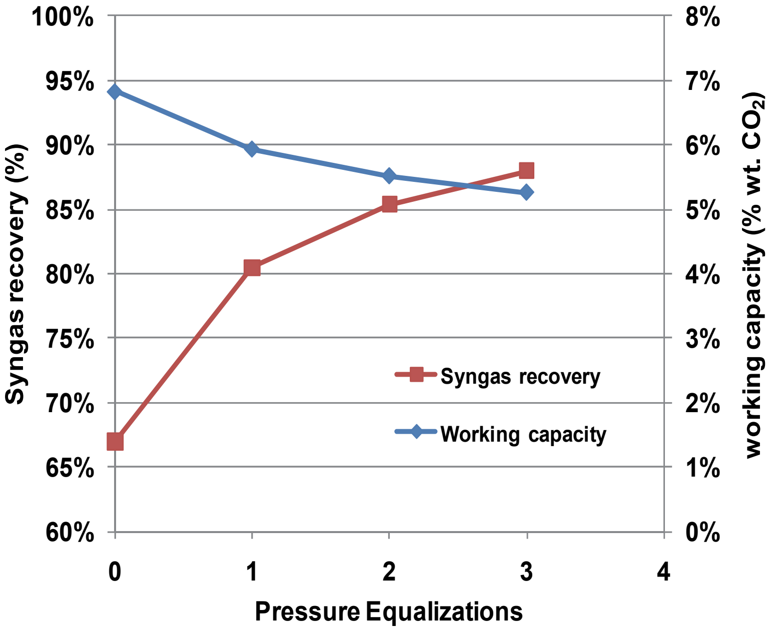


Figure 47. Optimization of number of pressure equalizations.

We optimized the amount of steam purge needed at 150 psia to provide very high CO_2 recovery and H_2 purity. The results are shown in Figure 48. We found that a steam purge of higher than 2.4 reactor volumes provides only marginal improvements (i.e., there is a change in slope as the steam purge volume is increased beyond 2.4 reactor volumes for both syngas recovery and working capacity). This shows 2.4 reactor volume is optimal. We also observed that as long as the steam purge volume is kept constant the purge time did not have any impact on the sorbent working capacity or the synthesis gas recovery.

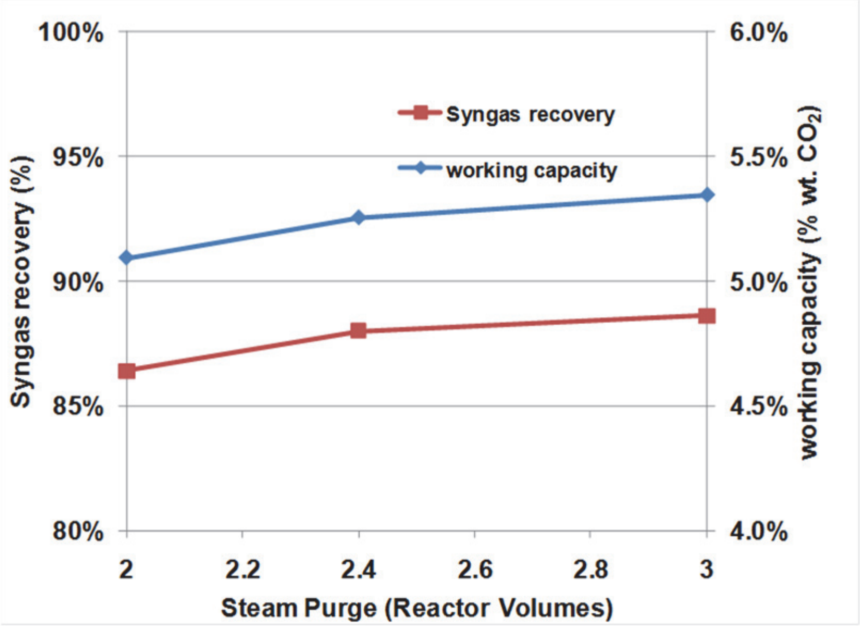


Figure 48. Optimization of steam purge volume.

Figure 49 shows the impact of desorption pressure on the synthesis gas recovery, working capacity and the steam consumption. The results indicate that lower desorption pressure provides lower steam consumption, lower working capacity and higher synthesis gas recovery. The lower desorption pressure will increase the CO_2 compression cost but lowers the CO_2 purification costs as the CO_2 product purity is higher (less of the syngas is contaminated with the CO_2).

We estimated the impact of desorption pressure by conducting a preliminary trade-off analysis against the steam consumption and the CO_2 compression power. Figure 50 shows the impact of desorption pressure on the overall auxiliary power consumption for steam purge and CO_2 compression. We observed that at higher pressures the CO_2 compression power decreases while the power loss due to steam consumption increases. We identified an optimum desorption pressure between 75-150 psia where the auxiliary power consumption is minimum.

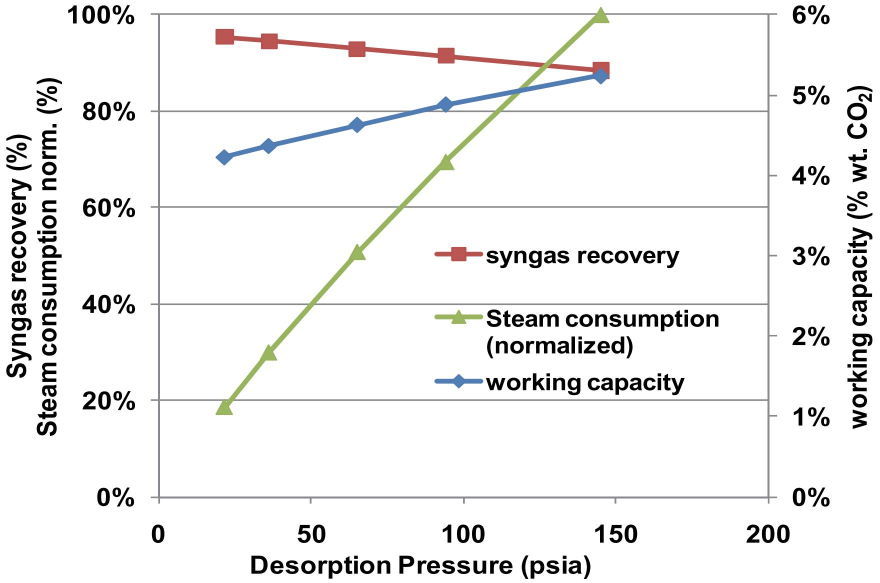


Figure 49. Optimization of desorption pressure or steam consumption.

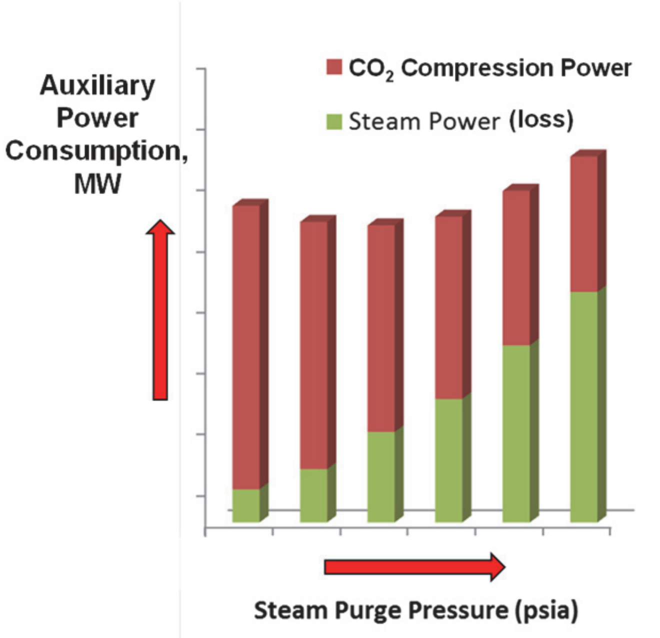


Figure 50. Auxiliary power consumption as a function of steam purge pressure.

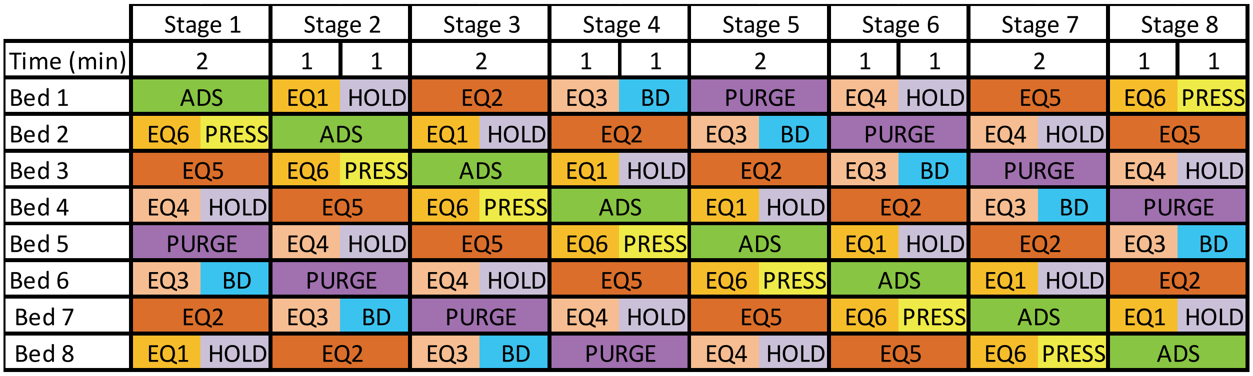
4.6.3 Optimization of Adsorption Temperature and Desorption Pressure for the Phillips 66 E-Gas^{TM} Gasifier Case

Based on the optimized PSA cycle scheme and parameters from the computer simulations, we designed the 8-bed PSA cycle scheme with three pressure equalizations. The PSA bed sequencing for the 8-bed PSA cycle scheme is shown in Table 13 and Figure 51. We initially designed the 8-bed PSA system to operate at an adsorption temperature of 240°C for the E-Gas^{TM} (Phillips 66) gasifier (i.e., at 77°C higher than the dew point of the synthesis gas with steam purge at two different pressures

Table 13. Steps in 8 bed PSA cycle scheme.

8- bed PSA Cycle Steps:

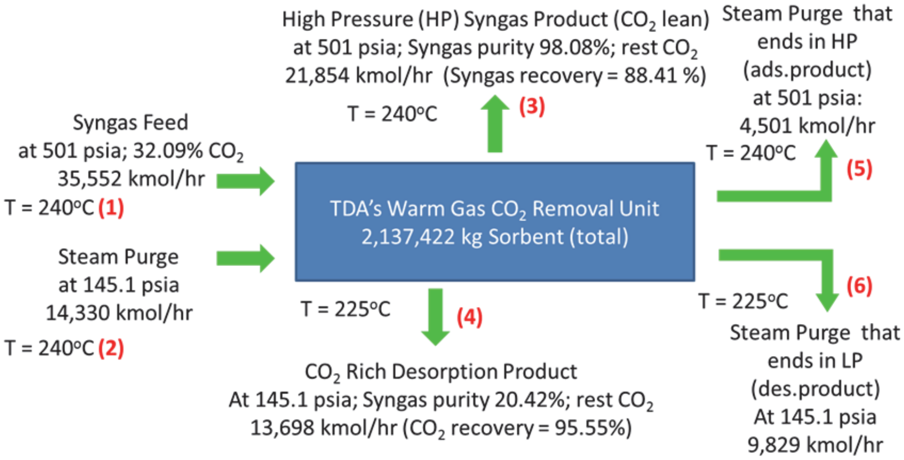
Step 1	Adsorption at 501 psia (ADS)
Step 2	Pressure Equalization to 420 psia (EQ1)
Step 3	Pressure Equalization to 340 psia (EQ2)
Step 4	Pressure Equalization to 260 psia (EQ3)
Step 5	Blowdown to 145.1 psia (BD)
Step 6	Steam Purge at 145.1 psia (PURGE)
Step 7	Pressure Equalization to 250 psia (EQ4)
Step 8	Pressure Equalization to 330 psia (EQ5)
Step 9	Pressure Equalization to 410 psia (EQ6)
Step 10	Product Pressurization to 501 psia (PRESS)



	Stage 1	Stage 2	Stage 3	Stage 4	Stage 5	Stage 6	Stage 7	Stage 8
Time (min)	2	1 1	2	1 1	2	1 1	2	1 1
Bed 1	ADS	EQ1 HOLD	EQ2	EQ3 BD	PURGE	EQ4 HOLD	EQ5	EQ6 PRESS
Bed 2	EQ6 PRESS	ADS	EQ1 HOLD	EQ2	EQ3 BD	PURGE	EQ4 HOLD	EQ5
Bed 3	EQ5	EQ6 PRESS	ADS	EQ1 HOLD	EQ2	EQ3 BD	PURGE	EQ4 HOLD
Bed 4	EQ4 HOLD	EQ5	EQ6 PRESS	ADS	EQ1 HOLD	EQ2	EQ3 BD	PURGE
Bed 5	PURGE	EQ4 HOLD	EQ5	EQ6 PRESS	ADS	EQ1 HOLD	EQ2	EQ3 BD
Bed 6	EQ3 BD	PURGE	EQ4 HOLD	EQ5	EQ6 PRESS	ADS	EQ1 HOLD	EQ2
Bed 7	EQ2	EQ3 BD	PURGE	EQ4 HOLD	EQ5	EQ6 PRESS	ADS	EQ1 HOLD
Bed 8	EQ1 HOLD	EQ2	EQ3 BD	PURGE	EQ4 HOLD	EQ5	EQ6 PRESS	ADS

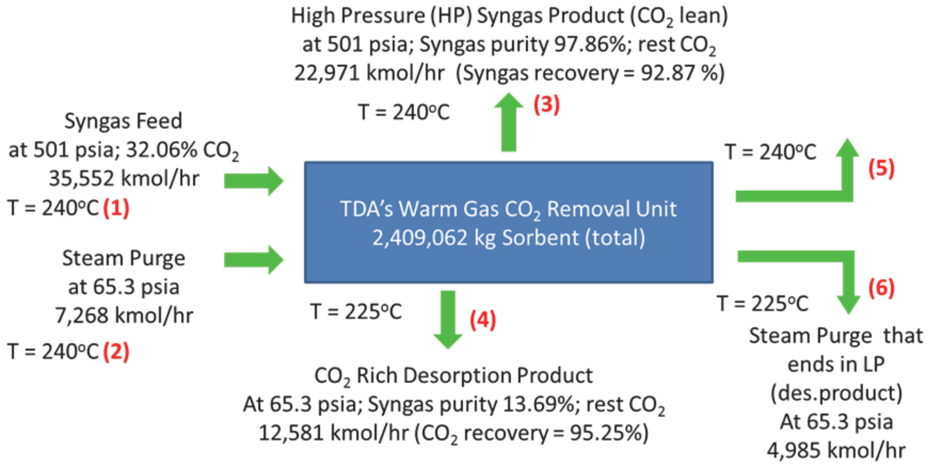
Figure 51. 8-bed PSA cycle sequence with three equalizations.

145 and 65 psia). The stream summaries around the high temperature PSA system are then calculated based on the simulation results for these two cases and were provided to UCI for inclusion into the Aspen Plus simulation to see the impact of the steam pressure on the net plant efficiency. Figure 52 and Figure 53 shows the simplified block diagram for these two cases. Aspen Plus process simulation results from UCI indicated that steam purge at 145.1 psia provided slightly better net plant efficiency than the low pressure case.



Note: 1. Streams (3) & (5) will be coming mixed together at high pressure
2. Streams (4) & (6) will be coming mixed together at low pressure

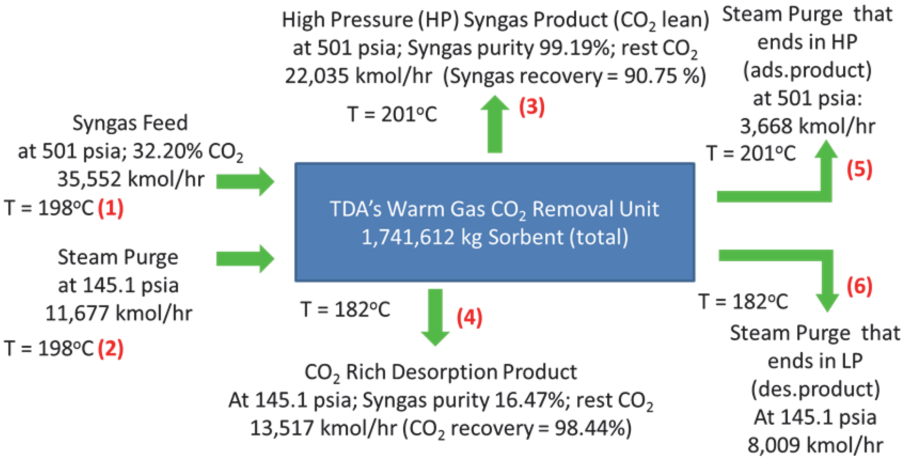
Figure 52. Simplified block diagram for the high temperature PSA system (TDA's Warm gas CO_2 removal unit) operating at an adsorption temperature of 240°C and a steam purge pressure of 145.1 psia.



Note: 1. Streams (3) & (5) will be coming mixed together at high pressure
2. Streams (4) & (6) will be coming mixed together at low pressure

Figure 53. Simplified block diagram for the high temperature PSA system (TDA's Warm gas CO_2 removal unit) operating at an adsorption temperature of 240°C and a steam purge pressure of 65.3 psia.

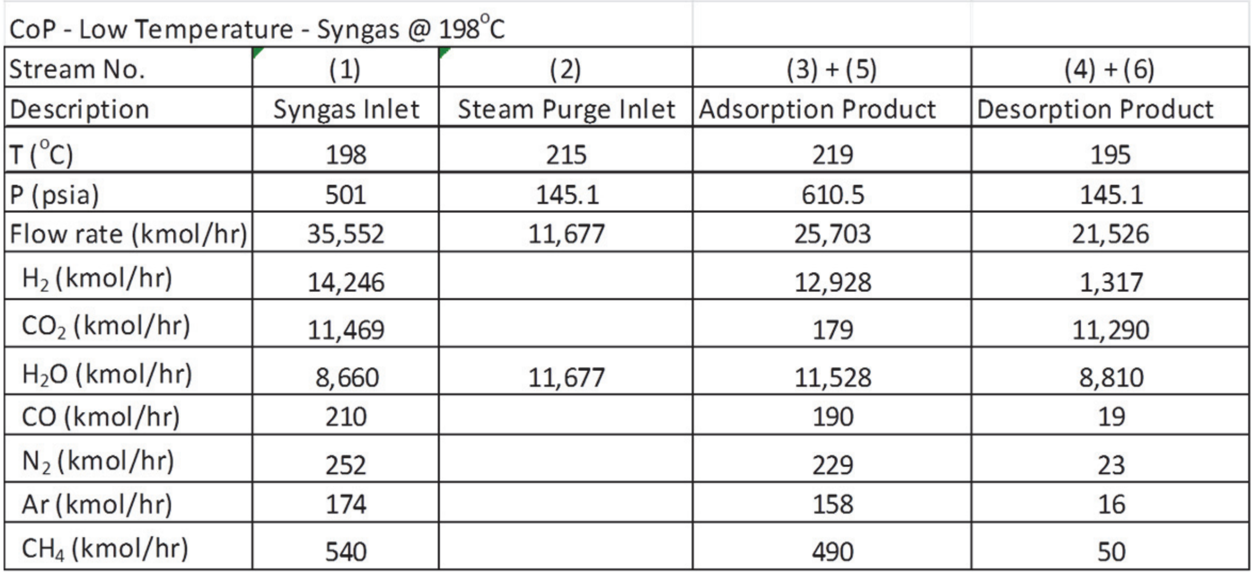
We then modified the PSA system design to carry out adsorption at a lower temperature (still above the dew point of the syngas) 198°C (35°C above the dew point of the syngas). This low temperature design slightly improved the net plant efficiency of the IGCC plant while significantly reducing the amount of sorbent needed from 2,409 tonnes to 1,741.6 tonnes. Simplified block diagram of the PSA system operating at an adsorption temperature of 198°C is provided in Figure 54 while the stream summary data is provided in Table 14.



Note: 1. Streams (3) & (5) will be coming mixed together at high pressure
2. Streams (4) & (6) will be coming mixed together at low pressure

Figure 54. Simplified block diagram for the high temperature PSA system (TDA's Warm gas CO_2 removal unit) operating at an adsorption temperature of 198°C and a steam purge pressure of 145.1 psia.

Table 14. Stream summary data for high temperature PSA system operating at 198°C for a Philips 66 E-Gas^{TM} gasifier.

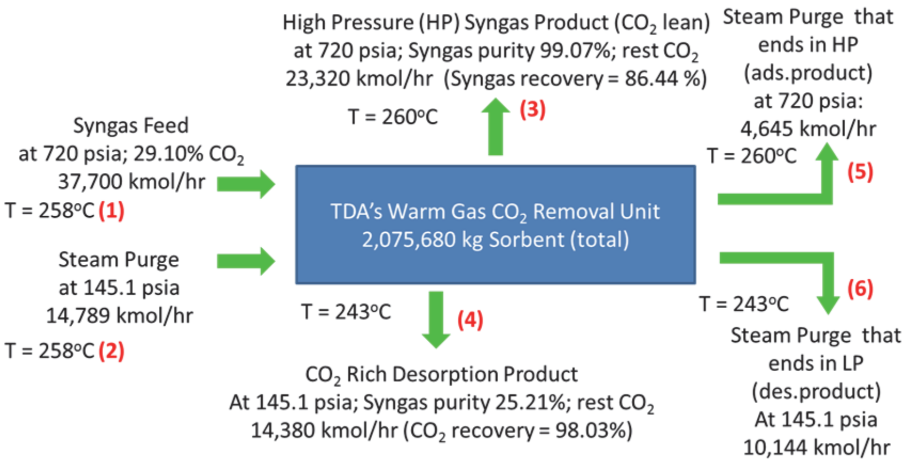


CoP - Low Temperature - Syngas @ 198°C

Stream No.	(1)	(2)	(3) + (5)	(4) + (6)
Description	Syngas Inlet	Steam Purge Inlet	Adsorption Product	Desorption Product
T (°C)	198	215	219	195
P (psia)	501	145.1	610.5	145.1
Flow rate (kmol/hr)	35,552	11,677	25,703	21,526
H_2 (kmol/hr)	14,246		12,928	1,317
CO_2 (kmol/hr)	11,469		179	11,290
H_2O (kmol/hr)	8,660	11,677	11,528	8,810
CO (kmol/hr)	210		190	19
N_2 (kmol/hr)	252		229	23
Ar (kmol/hr)	174		158	16
CH_4 (kmol/hr)	540		490	50

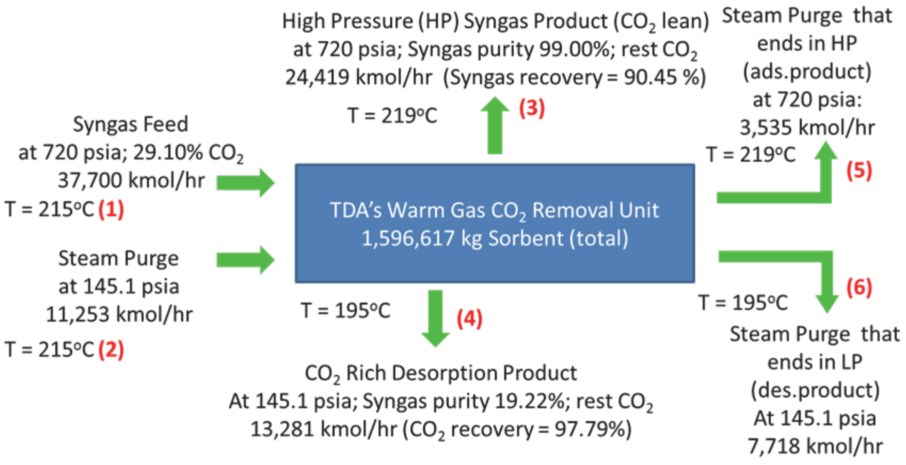
4.6.4 Optimization of Adsorption Temperature for the GE Gasifier Case

We then modified the PSA system design to carry out simulations for GE gasifier at two different adsorption temperatures 258 and 215°C (76 and 33°C above the dew point of the synthesis gas). For the GE gasifier low temperature design also led to slightly improvement in the net plant efficiency of the IGCC plant while significantly reducing the amount of sorbent needed from 2,075.7 tonnes to 1,596.6 tonnes. Simplified block diagrams of the PSA system operating at adsorption temperatures of 258 and 215°C are provided in Figure 55 and Figure 56, respectively. The stream summary data for the 215°C case is provided in Table 15.



Note: 1. Streams (3) & (5) will be coming mixed together at high pressure
2. Streams (4) & (6) will be coming mixed together at low pressure

Figure 55. Simplified block diagram for the high temperature PSA system (TDA's Warm gas CO_2 removal unit) operating at an adsorption temperature of 258°C and a steam purge pressure of 145.1 psia.



Note: 1. Streams (3) & (5) will be coming mixed together at high pressure
2. Streams (4) & (6) will be coming mixed together at low pressure

Figure 56. Simplified block diagram for the high temperature PSA system (TDA's Warm gas CO_2 removal unit) operating at an adsorption temperature of 215°C and a steam purge pressure of 145.1 psia.

Table 15. Stream summary data for high temperature PSA system operating at 215°C for a GE gasifier.

GE gasifier - Case 1 - Syngas @ 215°C				
Stream No.	(1)	(2)	(3) + (5)	(4) + (6)
Description	Syngas Inlet	Steam Purge Inlet	Adsorption Product	Desorption Product
T (°C)	215	215	219	195
P (psia)	720	145.1	720	145.1
Flow rate (kmol/hr)	37,700	11,253	27,954	20,999
H_2 (kmol/hr)	15,532		14,049	1,484
CO_2 (kmol/hr)	10,971		243	10,728
H_2O (kmol/hr)	10,518	11,253	13,049	8,723
CO (kmol/hr)	264		239	25
N_2 (kmol/hr)	188		170	18
Ar (kmol/hr)	188		170	18
CH_4 (kmol/hr)	38		34	4

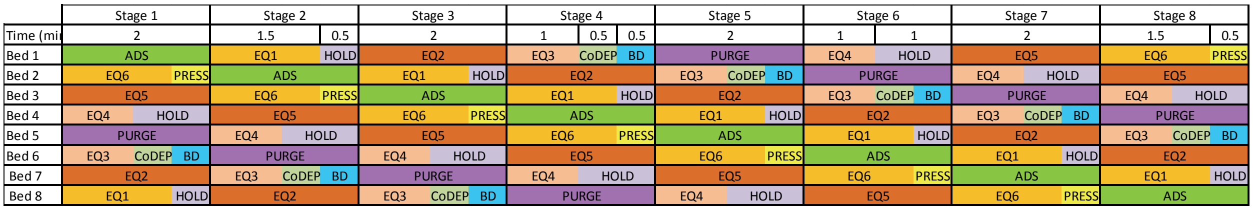
4.6.5 Design Improvements

In the DOE study for the case of Selexol^{TM} scrubbing based pre-combustion carbon capture, there is no additional cost accounted to purify CO_2, and the CO_2 stream stream purity is assumed to be 100%. However, it is to be noted that there will still be some Ar, N_2 mixed in with CO_2 based on their small but finite solubility of these compounds in the solvent at the regeneration conditions. In the early design of the TDA's CO_2 capture system, we selected a cryogenic purification system operating downstream of the CO_2 separation system to remove any H_2, CO, CH_4, N_2 and Ar impurities from the CO_2. Although the cryogenic separation was very effective for achieving very high CO_2 purity (reducing the overall contaminant concentration to less than 20 ppmv), it contributed to the higher cost for the purification process. In order to reduce this purification cost and eliminate the need for the cryogenic purification process to remove non-CO_2 synthesis gas compounds, we investigated multiple design options. In each one of these cases, in order to decrease the flow of these impurities to the purification system, we explored new PSA cycle schemes such as back purging with the product CO_2 and a co-current depressurization to recover more of the synthesis gas trapped in the pores of the sorbent to improve the CO_2 purity. The computer simulation results indicated that co-current depressurization is a better option than CO_2 purge. Hence, we designed the improved PSA cycle scheme with co-current depressurization for two different desorption pressures 145.1 psia and 65.3 psia.

Based on the optimized PSA cycle scheme and the parameters from the computer simulations we designed the 8-bed PSA cycle scheme with three pressure equalizations. The PSA bed sequencing for the 8-bed PSA cycle scheme is shown in Table 16 and Figure 57.

Table 16. Steps in the 8 bed PSA cycle scheme with co-current depressurization.

Step 1	Adsorption at 501 psia (ADS)
Step 2	Pressure Equalization to 420 psia (EQ1)
Step 3	Pressure Equalization to 340 psia (EQ2)
Step 4	Pressure Equalization to 260 psia (EQ3)
Step 5	Co-current Depressurization to 147 psia (CoDEP)
Step 6	Blowdown to 145.1 psia (BD)
Step 7	Steam Purge at 145.1 psia (PURGE)
Step 8	Pressure Equalization to 250 psia (EQ4)
Step 9	Pressure Equalization to 330 psia (EQ5)
Step 10	Pressure Equalization to 410 psia (EQ6)
Step 11	Product Pressurization to 501 psia (PRESS)



	Stage 1	Stage 2	Stage 3	Stage 4	Stage 5	Stage 6	Stage 7	Stage 8
Time (mi	2	1.5 0.5	2	1 0.5 0.5	2	1 1	2	1.5 0.5
Bed 1	ADS	EQ1 HOLD	EQ2	EQ3 CoDEP BD	PURGE	EQ4 HOLD	EQ5	EQ6 PRESS
Bed 2	EQ6 PRESS	ADS	EQ1 HOLD	EQ2	EQ3 CoDEP BD	PURGE	EQ4 HOLD	EQ5
Bed 3	EQ5	EQ6 PRESS	ADS	EQ1 HOLD	EQ2	EQ3 CoDEP BD	PURGE	EQ4 HOLD
Bed 4	EQ4 HOLD	EQ5	EQ6 PRESS	ADS	EQ1 HOLD	EQ2	EQ3 CoDEP BD	PURGE
Bed 5	PURGE	EQ4 HOLD	EQ5	EQ6 PRESS	ADS	EQ1 HOLD	EQ2	EQ3 CoDEP BD
Bed 6	EQ3 CoDEP BD	PURGE	EQ4 HOLD	EQ5	EQ6 PRESS	ADS	EQ1 HOLD	EQ2
Bed 7	EQ2	EQ3 CoDEP BD	PURGE	EQ4 HOLD	EQ5	EQ6 PRESS	ADS	EQ1 HOLD
Bed 8	EQ1 HOLD	EQ2	EQ3 CoDEP BD	PURGE	EQ4 HOLD	EQ5	EQ6 PRESS	ADS

Figure 57. 8-bed PSA cycle sequence with three equalizations and co-current depressurization.

Simplified block diagram of TDA's warm gas CO_2 removal unit (a high temperature PSA system) operating on an improved 8-bed PSA cycle scheme with three pressure equalizations and a co-current depressurization is shown in Figure 58. The various streams entering and leaving the TDA system are shown in Table 17.

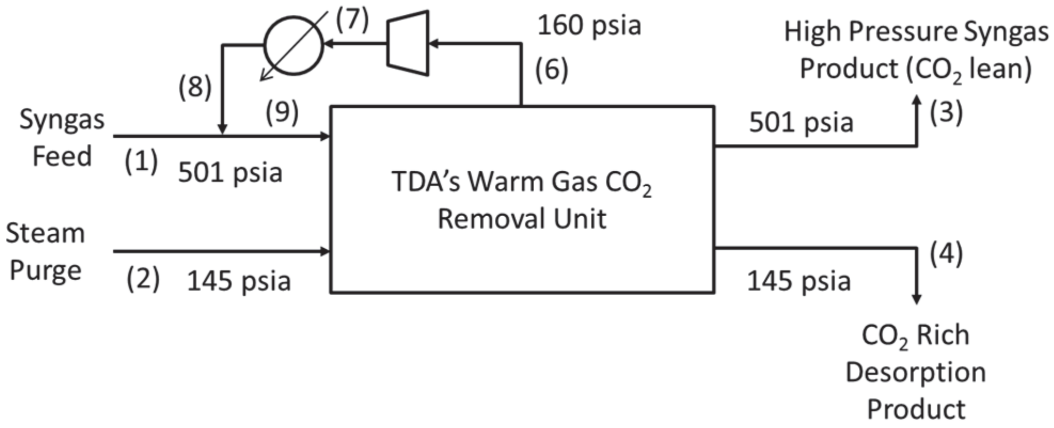


Figure 58. Simplified block diagram of TDA's CO_2 removal unit, a high temperatures PSA system containing 3 trains of 8-beds operating on an improved 8-bed PSA cycle scheme with three pressure equalizations and a co-current depressurization.

Table 17. Stream summary data for the improved high temperature PSA system operating at 198°C for a Phillips 66 (E-Gas™) gasifier.

Sorbent needed 2,061,893 kg

Stream No.	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)	(9)
Description	Syngas Inlet	Steam Purge Inlet	Syngas Product	CO_2 Product	CO_2 Purge in	Cocurrent depress out	Recycle compressor	Recycle	Bed Inlet
T (°C)	198	205	203	186	198	228	392	198	198
P (psia)	501	145.1	501	145.1	237	146.6	501	501	501
Flow rate (kmol/hr)	35,552	14,324	23,511	21,866	0	3,686	3,686	3,686	39,238
H2 (kmol/hr)	14,246		13,798	448	0	1,054	1,054	1,054	15,300
CO2 (kmol/hr)	11,469		185	11,285	0	1,904	1,904	1,904	13,373
H2O (kmol/hr)	8,660	14,324	12,888	10,097	0	641	641	641	9,301
CO (kmol/hr)	210		203	7	0	16	16	16	225
N2 (kmol/hr)	252		244	8	0	19	19	19	271
Ar (kmol/hr)	174		169	5	0	13	13	13	187
CH4 (kmol/hr)	540		523	17	0	40	40	40	580

TDA provided the composition, pressure, and temperature of these streams (Table 17) along with the bed sizing factor for sizing the sorbent requirement (182.777 kg sorbent/kmol CO_2 in product per h) to UCI. These flows were adjusted to fully load the gas turbines and UCI provided TDA with the final sorbent requirement.

4.6.6 High Temperature PSA System Design

For the final (and optimized) design case we calculated the amount of sorbent needed as 2,034 tonnes. Table 18 summarizes the bed sizing information. We designed the sorbent beds to have a L/D of 4 with three trains of 8-bed PSA systems containing a total of 24 beds that go through the PSA cycle sequence shown in Figure 57.

Table 18. TDA's Warm gas CO_2 removal unit (high temperature PSA system) bed sizing summary.

Sorbent Bed Sizing	
Syngas Flowrate on Total Plant Basis (already scaled up), kmol/hr	39,692
Sorbent Properties	
Sorbent Needed, kg / (kmol CO_2 in product/hr)	182.78
Sorbent bulk density, kg/L	0.42
Sorbent bulk density, lb/ft^3	26.22
CO_2 in Product, kmol/hr	11,128
Total Sorbent Needed, kg	2,033,936
Total Sorbent Needed, lb	4,484,056
Total Sorbent Volume, ft^3	171,019
Sorbent Volume per Reactor, ft^3	7,125.78
Reactor Info	
Number of Trains of 8-Beds Each	3
Total No. of Reactors	24
Reactor Size	
Aspect Ratio (Length/Diameter)	4
Diameter, ft	13
Length (T/T), ft	52
Operating Temperature, C	198
Operating Pressure, barA	33.82
Design Temperature, F	438
Design Pressure, psig	523
Single Vessel Basis	
Filling	
Bulk Density. lb/ft^3	26.22
Volume per Reactor, ft^3	6,902
lb per Reactor	180,970

Figure 59 shows a 3-D layout of the three-train of 8-bed PSA units with the PSA beds laid down in horizontal orientation. The selected design (i.e., horizontal orientation) will allow us to use smaller diameter reactors which ensures lower costs and easy transportation of the reactors to the application site (i.e., 13 ft reactors can be transported by rail and also can be trucked). Further design options were also considered with multiple inlets and outlets per bed to distribute the flow evenly. A cross section of the horizontally laid out bed with the sorbent loaded in it is shown in Figure 60.

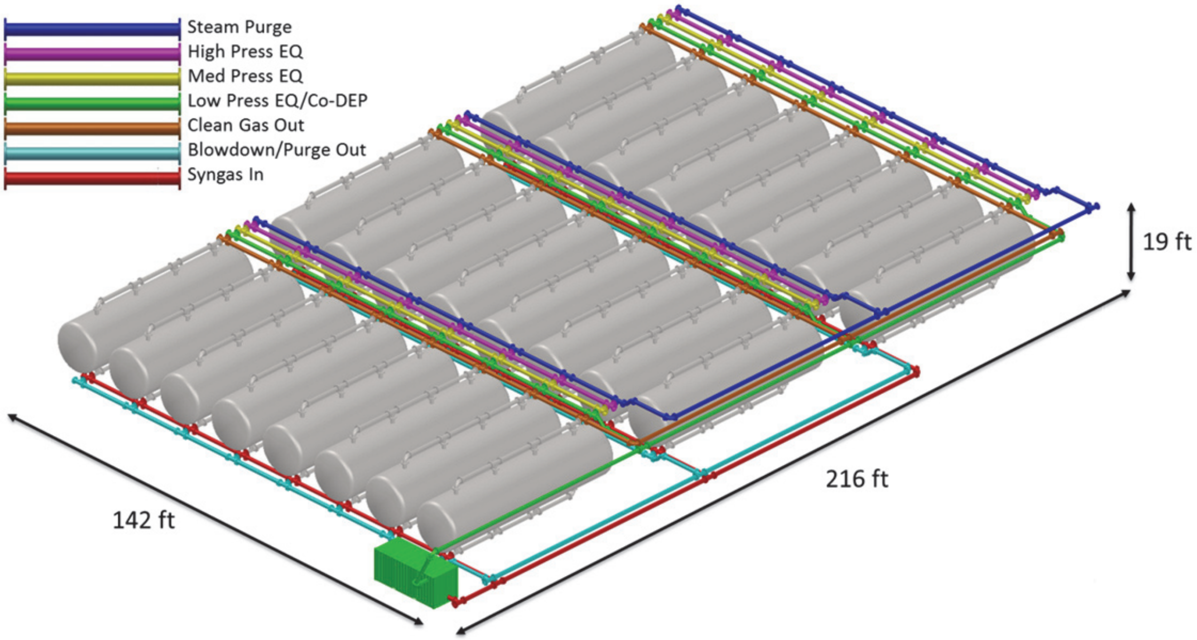


Figure 59. 3-D layout of TDA's warm gas CO_2 removal unit in horizontal orientation (three trains of 8-bed high temperature PSA system) operating on a PSA cycle sequence with three equalizations and co-current depressurization.

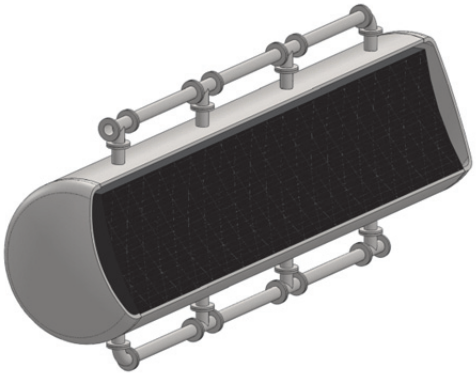


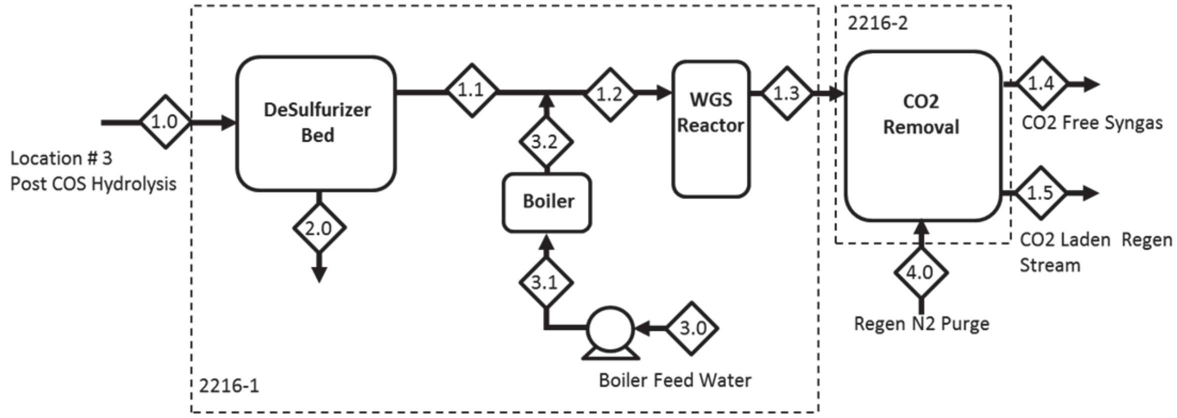
Figure 60. Cross section of the 3-D layout of horizontally laid PSA bed with sorbent loaded in it.

4.7 Task 7. Design of Prototype Test Unit

4.7.1 Process Flow Diagram and Stream Summaries for the Field Test Unit

In this task, TDA designed field demonstration test unit to evaluate the performance of its sorbent using actual synthesis gas. The tests at Wabash and Wilsonville were carried out in Task 9. The test unit consists of two parts: (1) the Gas Conditioning Unit and (2) the high temperature PSA-based CO_2 Separation Unit. The primary function of the synthesis gas conditioning unit was to adjust the concentration and purity of the synthesis gas. Because it is setup as a test site, the NCCC has all the capabilities of shifting the synthesis gas (converting the CO into CO_2 via water-gas-shift reaction), the Wabash River IGCC plant had no such capabilities. As a result in addition to the CO_2 separation unit, TDA designed and constructed a synthesis gas conditioning system with three major components; bulk desulfurization system, steam injector and a single stage water-gas-shift reactor.

As shown in Figure 61, synthesis gas (Stream 1.0) enters the Gas Conditioning Unit (GCU) where it flows through a sulfur removal bed to eliminate sulfur bearing contaminants (i.e., H_2S, COS). Next, the gas is mixed with steam produced by the GCU and enters a reactor containing water gas shift catalyst, where the CO-rich synthesis gas is converted to CO_2 and H_2 before entering the PSA Unit. Once in the PSA Unit, the gas flows through one of the four sorbent containing packed beds while the other three beds are either depressurizing or being regenerated using a nitrogen stream. Two NOVA fuel gas analyzers are in place for online measurement of the concentrations of CO_2, CO and H_2 in both the synthesis gas exit stream as well as the regeneration off-gas stream.

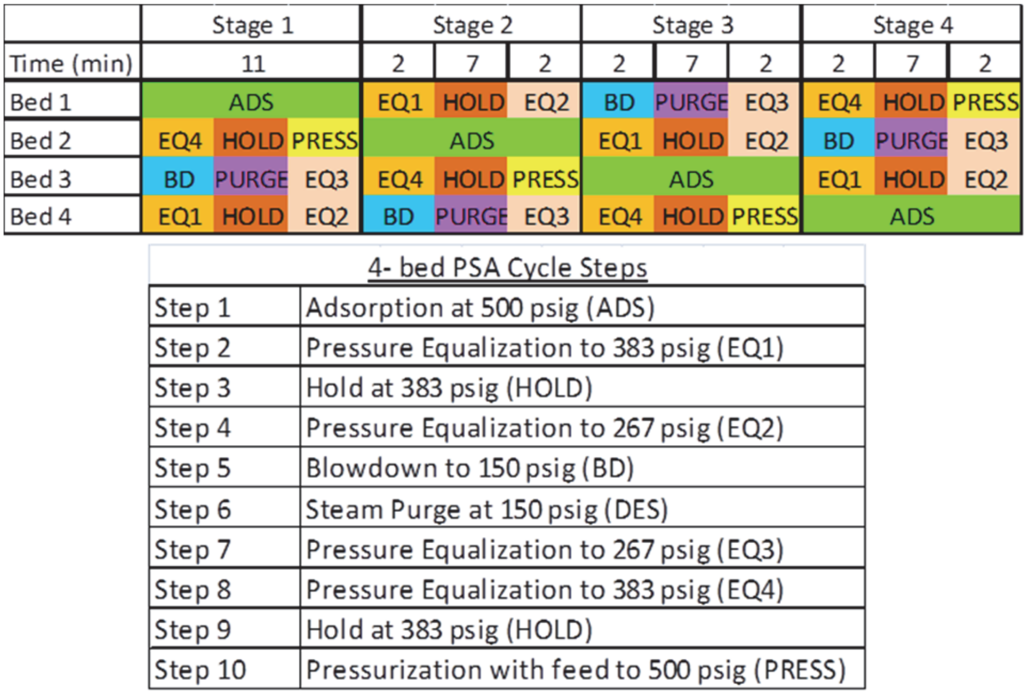


Average Flows												
Stream		1.0	1.1	1.2	1.3	1.4	1.5	2.0	3.0	3.1	3.2	4.0
gmol/hr												
H2O		7.5	7.5	27.3	14.3	14.3	0.0	0.0	19.8	19.8	19.8	0.0
H2		9.1	9.1	9.1	22.1	21.5	0.7	0.0	0.0	0.0	0.0	0.0
H2S		0.5	0.0	0.0	0.0	0.0	0.0	0.5	0.0	0.0	0.0	0.0
CO		13.7	13.7	13.7	0.6	0.6	0.0	0.0	0.0	0.0	0.0	0.0
CO2		5.2	5.2	5.2	18.3	0.2	18.1	0.0	0.0	0.0	0.0	0.0
N2		0.9	0.9	0.9	0.9	0.0	81.2	0.0	0.0	0.0	0.0	80.3
Other		1.0	1.0	1.0	1.0	1.0	0.0	0.0	0.0	0.0	0.0	0.0
Total		**37.9**	**37.4**	**57.2**	**57.2**	**37.5**	**100.0**	**0.5**	**19.8**	**19.8**	**19.8**	**80.3**
Temp	C	193.8	250.0	293.8	350.0	110.0	85.0		20.0	20.0	350.0	25.0
Press	PSIG	357.0	356.0	356.0	356.0	1.5	2.0		50.0	450.0	450.0	90.0

Figure 61. The Process Flow Diagram and stream summary for the field test unit.

The PSA-based CO_2 removal system consists of four packed beds filled with TDA's CO_2 sorbent. Although the slipstream test unit did not use the 8-bed configuration selected for the final design, the selected configuration was more than adequate for demonstrating all critical aspects of the concept within the budget and schedule constraints of this project. The 4-bed configuration allowed continuous CO_2 removal and product flow. The bed and valve layout was designed around the cycle sequence shown in Figure 62. Each bed has three valves at the inlet section and two valves at the outlet to allow for continuous product flow during regeneration and depressurization.

The 4-bed configuration also allow us to carry out pressure equalizations (by allowing the low pressure regenerated bed to partially pressurize with the high pressure bed that has just finished with adsorption step) and provided decent hydrogen recovery.



	Stage 1			Stage 2			Stage 3			Stage 4		
Time (min)	11			2	7	2	2	7	2	2	7	2
Bed 1	ADS			EQ1	HOLD	EQ2	BD	PURGE	EQ3	EQ4	HOLD	PRESS
Bed 2	EQ4	HOLD	PRESS	ADS			EQ1	HOLD	EQ2	BD	PURGE	EQ3
Bed 3	BD	PURGE	EQ3	EQ4	HOLD	PRESS	ADS			EQ1	HOLD	EQ2
Bed 4	EQ1	HOLD	EQ2	BD	PURGE	EQ3	EQ4	HOLD	PRESS	ADS		

4- bed PSA Cycle Steps	
Step 1	Adsorption at 500 psig (ADS)
Step 2	Pressure Equalization to 383 psig (EQ1)
Step 3	Hold at 383 psig (HOLD)
Step 4	Pressure Equalization to 267 psig (EQ2)
Step 5	Blowdown to 150 psig (BD)
Step 6	Steam Purge at 150 psig (DES)
Step 7	Pressure Equalization to 267 psig (EQ3)
Step 8	Pressure Equalization to 383 psig (EQ4)
Step 9	Hold at 383 psig (HOLD)
Step 10	Pressurization with feed to 500 psig (PRESS)

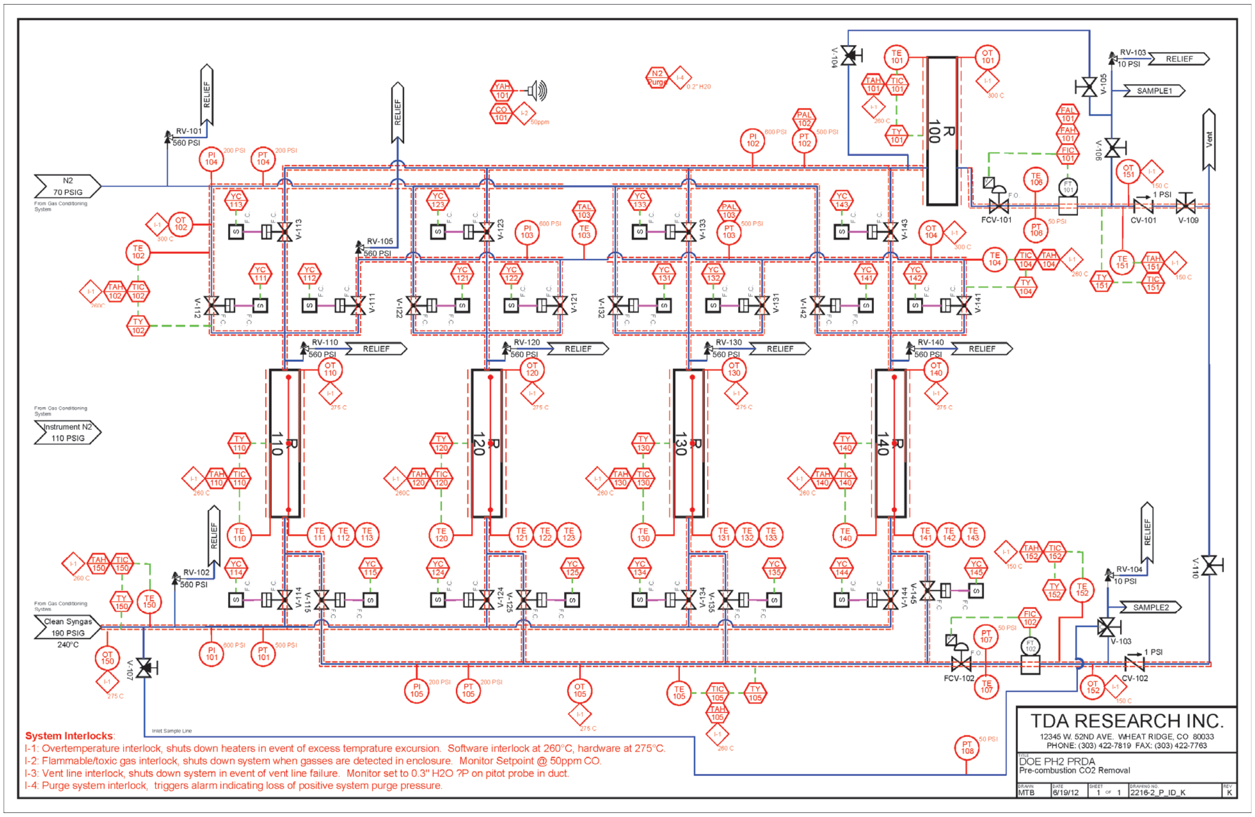
Figure 62. 4-Bed PSA Cycling Sequence.

4.7.2 Detailed Design of the Test Unit

Figure 63 shows the P&ID of the PSA-based CO_2 Removal Unit. The shifted and sulfur-free synthesis gas enters the system and is diverted to any one of the four beds depending on the orientation of the valves located at the bottom of each reactor vessel. All the gas lines in the system are heat traced and all the valves are located in metal boxes that are heated above the dew point of the gas in order to prevent water and hydrocarbons from condensing in the manifolds or over the sorbent bed. To maintain the bed temperature, each reactor is wrapped in a heated and insulated jacket and the vessel surface temperature is precisely controlled. After adsorption, the CO_2-free synthesis gas flows into the accumulator vessel (R-100). The accumulator is sized approximately four times that of the volume of the reactors. Its purpose is to pressurize the reactors with CO_2 free synthesis gas once regeneration and equalization has been completed. The gas flow out of the system is controlled by a Badger Research Control Valve, while flow rate is measured by a turbine flow meter.

Once a bed is close to its breakthrough for CO_2, the next step is to rapidly “blow down” or depressurize the bed to help drive off the CO_2 that is adsorbed on the surface of the sorbent. Once the bed reaches approximately 75 psig, heated nitrogen is provided to the bed from the top down to continue to drive off the remaining CO_2 adsorbed on the surface of the sorbent (this was to simulate the steam purge during the regeneration step). A Badger Research Control Valve controls the flow of nitrogen along with the rate in which the bed depressurizes. A mass flow meter measures the amount of nitrogen that has flown through the bed. A turbine flow meter measures the combined outlet flow to help determine the amount of the CO_2 and H_2 that is in the regeneration stream.

Gas samples are taken for analysis at multiple points in the system. A sample stream is taken from the synthesis gas outlet and sent to a gas analyzer which is capable of determining the concentrations of CO_2, CO and H_2. Gas samples can also be taken before or after the accumulator in order to analyze the performance of each individual bed or the entire system. Finally, there is also a sample stream from the regeneration side and sent to one of the NOVA analyzer to determine the CO_2, CO and H_2 content in the recovered CO_2 stream.



System Interlocks:
I-1: Overtemperature interlock, shuts down heaters in event of excess temprature excursion. Software interlock at 260°C, hardware at 275°C.
I-2: Flammable/toxic gas interlock, shuts down system when gasses are detected in enclosure. Monitor Setpoint @ 50ppm CO.
I-3: Vent line interlock, shuts down system in event of vent line failure. Monitor set to 0.3" H2O ?P on pitot probe in duct.
I-4: Purge system interlock, triggers alarm indicating loss of positive system purge pressure.

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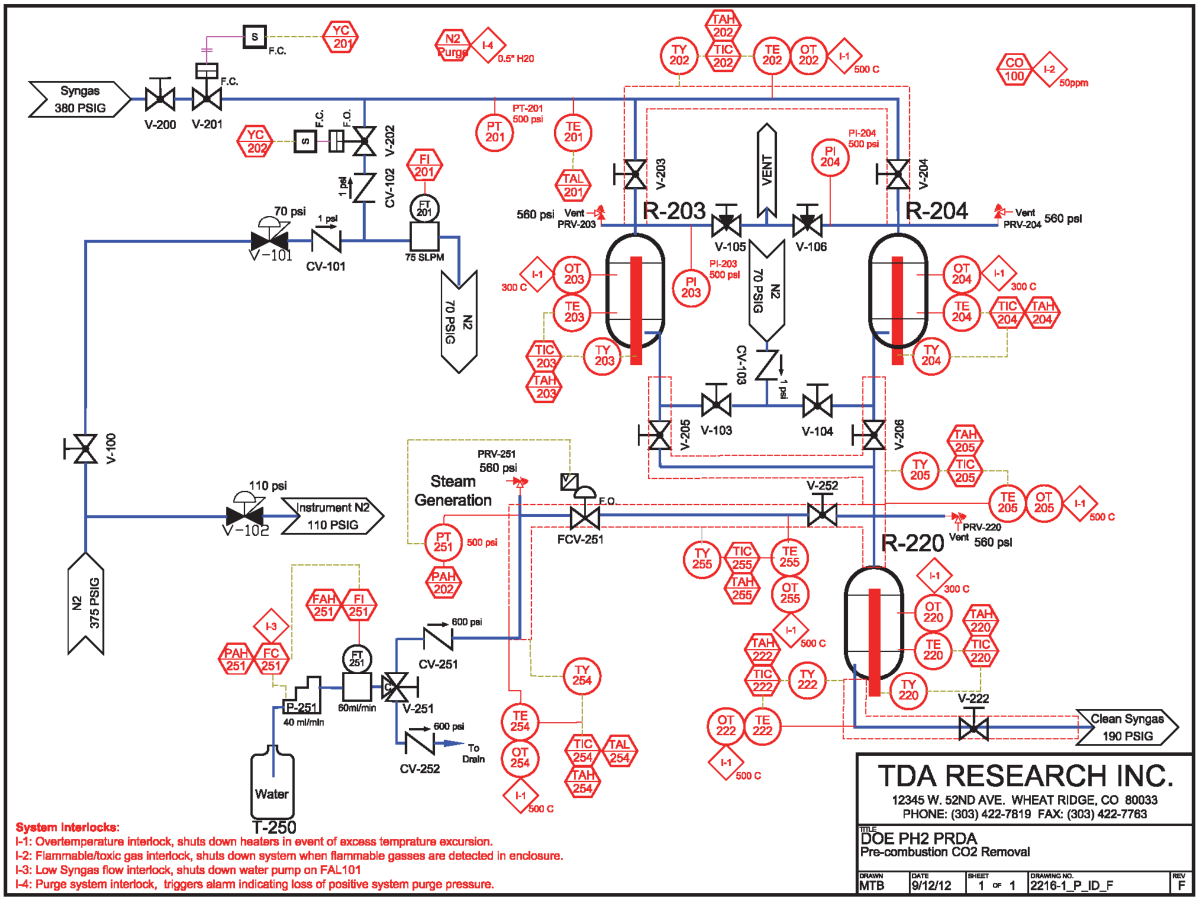
TITLE				
DOE PH2 PRDA Pre-combustion CO2 Removal				
DRAWN	DATE	SHEET	DRAWING NO.	REV
MTB	6/19/12	1 OF 1	2216-2_P_ID_K	K

Figure 63. The P&ID of the CO_2 Adsorption Unit.

Before entering the CO_2 adsorption unit, the synthesis gas is first processed by the gas conditioning unit. As shown in the P&ID (Figure 64), the synthesis gas first passes through either R-203 or R-204 vessels. Both reactors contain Actisorb® sorbent, a chemical sulfur removing sorbent commercially provided by SudChemie (now Clariant). Two identical reactors are used so that when one of the beds is fully saturated with sulfur, the gas flow is diverted to the other bed with minimal interruption to the operation. Both vessels are internally heated using cartridge heaters. After desulfurization, steam is added to the synthesis gas to raise the steam:carbon ratio in order to perform water gas shift reaction. Water is pumped into a heated ¼" line and is vaporized. A Badger Research Control Valve maintains slightly higher pressure in the steam generation line to ensure that the steam entering the syngas is completely vaporized. The synthesis gas and steam mixture then enters to R-220 vessel which contains the Shiftmax® 240 water-gas-shift catalyst (a highly common low temperature WGS catalyst provided SudChemie). This is a copper based catalyst (copper and zinc oxide on alumina) performs the water gas shift reaction:

$$CO + H_2O \rightarrow CO_2 + H_2$$

The remaining CO content in the outlet stream is continuously monitored to ensure that there is enough steam to not only shift the CO but to prevent coking of the catalyst. The shifted syngas then exits the gas condition unit via a heated gas line to the CO_2 adsorption unit.



System Interlocks:
I-1: Overtemperature interlock, shuts down heaters in event of excess temprature excursion.
I-2: Flammable/toxic gas interlock, shuts down system when flammable gasses are detected in enclosure.
I-3: Low Syngas flow interlock, shuts down water pump on FAL101
I-4: Purge system interlock, triggers alarm indicating loss of positive system purge pressure.

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Pre-combustion CO2 Removal

DRAWN	DATE	SHEET	DRAWING NO.	REV
MTB	9/12/12	1 OF 1	2216-1_P_ID_F	F

Figure 64. The P&ID of the Gas Conditioning Unit.

4.7.3 Safety Features

Before finalizing the Test Unit's P&IDs, a thorough Process Hazard Analysis (PHA) was performed in conjunction with engineers from ConocoPhillips (now Phillips 66) and the operators of the Wabash River IGCC Power Plant to ensure the all aspects of the Test Unit are safe for operation and complies with onsite rules and regulations. A safe operating procedure was also developed to ensure that safe operation of the test unit is always followed:

1. Both units are housed in NEMA enclosures which are purged with nitrogen in accordance with standard Class I Division II practices. Purge monitors measure the

pressure created by the purge and will shut the system down if there is insufficient flow to the enclosures.

2. Steam rated pressure relief valves are located throughout the entire test unit in order to prevent over-pressurization of the vessels and tubing. Pressure throughout the entire unit is monitored using a series of pressure transducers. Each pressure measurement is associated with a software interlock which will shut the system down if the measured pressure exceeds a predetermined set point.
3. Every heater in the system is monitored by two thermocouples. One is wired directly to an over-temperature module which will terminate power to all heaters if the measured temperature exceeds the set point on the module. The other thermocouple is wired to the Labview control software which is used for control. The software will shut the system down if the monitored temperature exceeds the predetermined set point.
4. Combustible gas monitors are located in both cabinets to monitor possible leakage inside the enclosures. A shutdown will be initiated if the amount of combustible gases exceeds a safe level within the enclosure.
5. All the fittings and tubing are rated well above the maximum limit of the apparatus for pressure and temperature ratings. All fittings used in high pressure and temperature regions are Swagelok® compression fittings. Each vessel is ASME rated for well above the temperatures and pressures of the system.

A similar Process Hazard Analysis and Safety Review were also carried out with the Southern Company engineers to qualify the unit at the NCCC site.

4.8 Task 8. System Analysis and Process Economic Analysis Evaluation

Under subcontract to TDA Research, the Advanced Power and Energy Program of the University of California, Irvine (UCI) provided assistance in the process design and developed system simulation models to assess the economic viability of TDA's high temperature PSA-based pre-combustion carbon capture (Warm Gas Cleanup) technology. The analysis also provided a thorough comparison against a conventional cold gas cleanup technology based on the Selexol™ physical solvent scrubbing (used for desulfurization and de-carbonization of the raw synthesis gas). Using the performance data for the CO_2 removal system provided by TDA (that consisted of the definition of all key process parameters including the gas composition, flow rate, temperature and pressure of each process stream entering and leaving the PSA system), UCI developed the overall process model with all sub-systems.

In the analysis, UCI investigated the integration of the high temperature PSA system with the Phillips 66 and General Electric (GE) gasification technologies. Various design cases were modeled using the same coal type (Illinois #6 bituminous coal). In these simulations UCI evaluated the use of different operating temperatures for the high temperature PSA system for both gasifier types.

- Operation at 198^{o}C and 240^{o}C i.e., 35^{o}C and 77^{o}C higher than the dew point of the synthesis gas, respectively for the system integrated with Phillips 66 gasifier
- Operation at 215^{o}C and 258^{o}C i.e., 33^{o}C and 76^{o}C higher than the dew point of the synthesis gas, respectively for the system integrated with GE gasifier

The analysis results suggested that the lower operating temperature in each gasifier case provided a lower CO_2 removal cost due to increased sorbent performance (sorbent achieves a higher CO_2 capacity at low temperatures maintaining all other operating parameters the same).

UCI also carried out simulations based on the use of two separate CO_2 purification and compression options:

- A cryogenic separation system, removing CO_2 as a liquid from the bottoms of a cryogenic distillation tower and pressurizing it further using liquid pumps
- A water removal/compressor-based purification/compression system that uses gas phase compression combined with liquid CO_2 pumping to deliver CO_2 as a supercritical fluid meeting the desired pipeline delivery requirement (similar to the approach used for the Cold Gas Cleanup Case).

We identified that the gas phase compression followed by liquid pumping is more economical than the cryogenic purification and compression process, while the latter approach provides higher CO_2 purity (reducing the concentration of Ar and N_2 to less than 20 ppmv). Since with the gas phase CO_2 compression option, the product CO_2 purity from TDA's high temperature PSA system matched the purity level that can be achieved by the Selexol^{TM} process, this option was selected for further analysis due to its lower capital and operating cost.

Finally, several cases have been analyzed to assess the impact of the regeneration pressure and the overall steam consumption during the reactor purge step on sorbent performance and on the overall plant cost. The results suggested that the 9.7 barA regeneration process (the pressure at which the CO_2 product gas is recovered) with 66,350 kg of steam consumption per cycle (at 150 psia and 205^{o}C) associated with 2.4 bed volumes of steam purge provided the lowest operating cost for the power plant.

A detailed description for the most efficient and cost effective pre-combustion carbon capture system design integrated with the Phillips 66 gasifier is presented in the following sections of this report. The complete analysis results for all other cases carried out throughout the course of the project are also included in the Appendix section.

4.8.1 Process Design

Figure 65 shows a simplified block diagram of an IGCC plant integrated with TDA's high temperature PSA-based CO_2 capture process. TDA's Warm gas CO_2 capture system is located downstream of a warm gas desulfurization system (we used the regenerable zinc-titanate-based syngas desulfurization system developed by the Research Triangle Institute). The regeneration off-gas from the desulfurization system is further treated in a sulfuric acid plant, ultimately converting all sulfur into a concentrated H_2SO_4 product. The desulfurized synthesis gas is then fed to the water-gas-shift system that converts the CO into H_2. Consistent with the DOE analysis, we maintained a H_2O:CO molar ratio of 2.0 at the inlet of the high temperature shift reactor (1 extra mole of H_2O than required by reaction stoichiometry). The synthesis gas from the water gas shift unit is sent to the CO_2 Capture system at a temperature slightly (at least 30^{o}C) above the dew point of the synthesis gas. This eliminates the need to cool the synthesis gas below its dew point using condensing heat exchangers.

The CO_2 capture block consists of a CO_2 separation system (the high temperature PSA system) and a purification/compression system (which further treats the CO_2 stream from the separation unit into a pure, pressurized CO_2 product that meets pipeline specification). As required by the FOA, TDA's high temperature PSA-based CO_2 separation system captures 90% of the carbon in the synthesis gas as CO_2 and produces a CO_2-lean synthesis gas that is sent to the gas turbine. Any gases trapped in the voids of the sorbent and the reactor ullage space are recovered at an intermediate pressure and recycled back to the synthesis gas feed to ensure

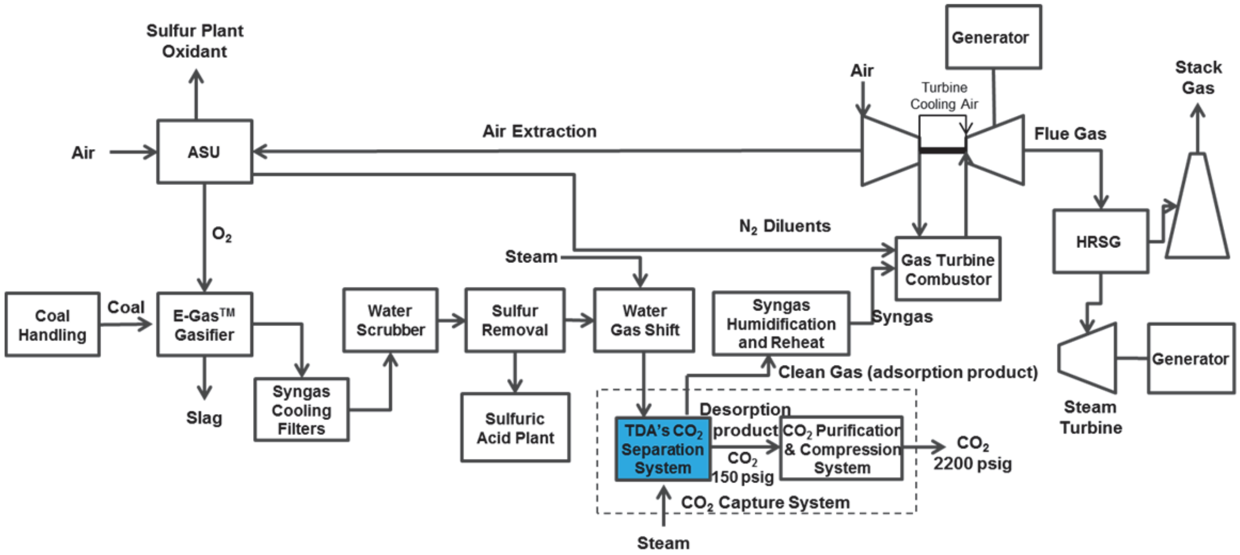


Figure 65. TDA's CO_2 Separation System Integrated with an IGCC plant.

high H_2 recovery in the CO_2 separation unit. A steam purge at lower pressure is used to fully regenerate the sorbent. A CO_2 rich stream primarily consisting of H_2O and CO_2 (along with some CO and H_2 impurities) is sent for further purification (either using cryogenic separation or a catalytic combustor to burn any residual syngas with oxygen) and compression to produce high purity CO_2 at 2,200 psig that can be sent for sequestration.

4.8.2 TDA's CO_2 Separation System

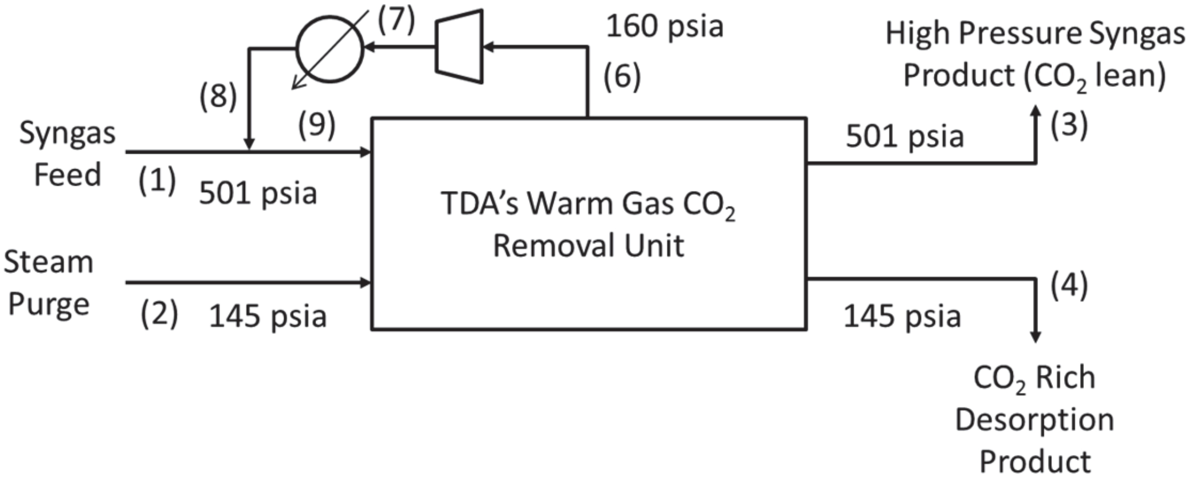


Figure 66. Simplified block diagram of TDA's CO_2 removal unit.

Simplified block diagram of TDA's warm gas CO_2 removal unit along with the various streams entering and leaving the system are shown in Figure 66. TDA provided the composition, pressure, and temperature of these streams along with the bed sizing factor for sizing the sorbent requirement and the 3-D layout for the system. TDA's warm gas CO_2 capture system consists of three trains of 8-bed PSA systems containing a total of 24 beds that go through the following PSA cycle steps. A 3-D layout of the TDA's warm gas CO_2 capture system provided by TDA is shown in Figure 67.

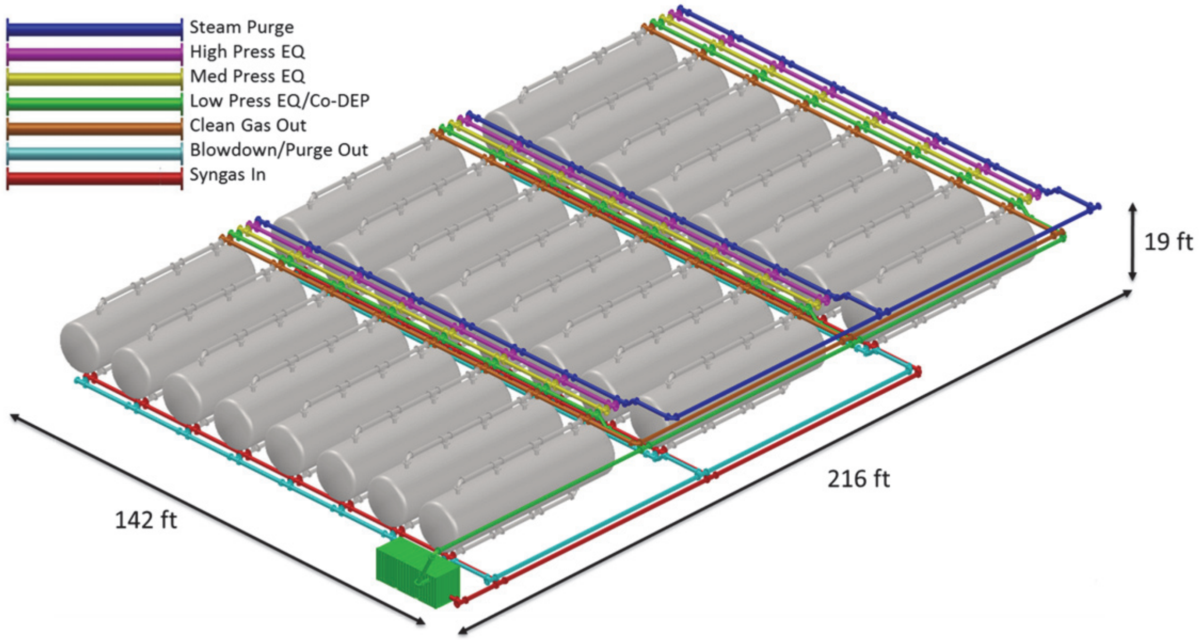


Figure 67. 3-D layout of TDA's warm gas CO_2 removal unit in horizontal orientation (three trains of 8-bed high temperature PSA system) operating on a PSA cycle sequence with three equalizations and co-current depressurization.

4.8.3 Process Design Basis and Methodology

Consistency has been maintained between the IGCC case developed under this study utilizing the TDA's high temperature PSA-based CO_2 capture process (**Warm Gas Cleanup Case**) with the Phillips 66 gasifier based IGCC plant with CO_2 capture utilizing current state-of-the-art syngas cleanup and CO_2 capture technology (**Cold Gas Cleanup Case**). The Cold Gas Cleanup Case is modeled so that it is consistent with the Case 4 in the updated DOE NETL study report DOE/NETL-2010/1397 titled, "Cost and Performance Baseline for Fossil Energy Plants," dated November 2010.

The CO_2 adsorption in the Warm gas Cleanup Case is carried out in a high temperature pressure swing adsorption (PSA) unit that uses TDA's low cost, high capacity CO_2 sorbent. The sorbent can also simultaneously remove the trace contaminants such as Hg and As that are present in the synthesis gas. Performance data for the CO_2 adsorption unit (with co-sorption of Hg) were provided by TDA and consisted of the definition of the streams (composition, flow rate, temperature and pressure) associated with this subsystem for a given raw syngas stream as estimated by UCI. Sizing basis for the CO_2 sorption vessels using the pressure swing adsorption (PSA) process was also provided by TDA.

Site Conditions

The plant is designed for the following mean ambient conditions (ISO conditions):

- Dry bulb temperature: 15°C
- Elevation: Mean Sea Level
- Relative Humidity: 60%

Mechanical draft cooling towers with 11°C temperature rise for the cooling water are used. The above ambient conditions correspond to a wet bulb temperature of 10.8°C. A cooling water supply temperature of 15.56°C (60°F) is used which correspond to a reasonable approach temperature to the wet bulb temperature.

Coal Feed

The characteristics of the coal feed are the same as that in the previous DOE/NETL study (DOE/NETL-2010/1397) and are shown in Table 19.

Table 19. Coal Data

Rank	**High-volatile Bituminous**	
Seam	**Illinois #6 (Herrin)**	
Sample	**St. Clair Co., IL**	
Proximate Analysis (wt. %)	**As Received**	**Dry**
Moisture	11.12	0
Ash	9.7	11
Volatile Matter*	34.99	39
Fixed Carbon	44.19	50
HHV		
kJ/kg	27,113	30,506
Btu/lb	11,666	13,126
LHV		
kJ/kg	26,151	29,544
Btu/lb	11,252	12,712
Ultimate Analysis (wt. %)	**As Received**	**Dry**
Carbon	63.75	71.72
Hydrogen	4.5	5.06
Nitrogen	1.25	1.41
Chlorine	0.29	0.33
Sulfur	2.51	2.82
Ash	9.7	10.91
Oxygen^{+}	6.88	7.75
Moisture	11.12	0
Total	100	100

* assumed all sulfur as volatile matter, + by difference

Carbon Dioxide Product Specification

The specifications of the CO_2 product are based on the typical purity obtained by the Selexol™ cold gas cleanup process which was used in the previously referenced DOE/NETL report without any further purification. Initial cases included in the Appendix were developed using 300 ppmv as the limit for N_2 and 10 ppmv as the limit for Ar. In order to meet these requirements an expensive (both in terms of cost and efficiency penalty) distillation unit was required for purification of the raw CO_2 produced by the PSA unit. These limits were later relaxed so that the Argon level in the CO_2 stream coming off the TDA's PSA unit is similar to what is typically obtained from a Selexol™ unit, i.e., without any further purification. This is done to be consistent with the DOE analysis referenced earlier (DOE/NETL-2010/1397). This eliminated the need for cryogenic distillation in the CO_2 purification process and was replaced with a catalytic combustor to burn any residual syngas present in the CO_2 stream with oxygen from the air separation unit (ASU). Table 20 summarizes the CO_2 specifications used for this analysis.

Table 20 CO_2 Pipeline Specification

Parameter Units	Parameter Value
Inlet Pressure MPa (psia)	15.3 (2,215)
Outlet Pressure MPa (psia)	10.4 (1,515)
Inlet Temperature °C (°F)	35 (95)
CO_2 Concentration, mole %	> 95
N_2 Concentration	not limited
O_2 Concentration ppmv	< 40
Ar Concentration	not limited
H_2O Concentration ppmv	< 150

Makeup Water

Makeup water as required by the plant is assumed to be fresh water without any "unusual" treatment requirements.

Property Packages

The performance of the plant was estimated using the Aspen Plus Simulator software, and the following property packages were used to model the cycle:

- PR-BM (Peng-Robinson equation of state with Boston-Mathias modifications) for most plant section systems except as follows
- ElecNRTL (Electrolyte Non-random Two Liquid model, also known as the Chen electrolyte model, formulated for modeling aqueous electrolytes) for sour water systems
- STEAM-TA (ASME 1967 Steam Table Corrections) for the water and steam.

4.8.4 Process Descriptions

The Calibration Case was first modeled in Aspen Plus® and the overall thermal performance of the plant was compared to that of the DOE/NETL study Case 4 in order to validate the Aspen Plus® model developed for the entire IGCC system.

Cold Gas Cleanup

The IGCC plant employing the cold gas cleanup and CO_2 capture technology consists of the following plant subsystems:

- Air Separation Unit (ASU)
- Coal Feed Preparation
- Gasification (based on Phillips 66 technology)
- High Temperature Syngas Cooling and Scrubbing
- Sour Shifting and Cold Gas Heat Recovery

- Syngas Desulfurization and Decarbonization using a two-stage Selexol™ process
- Claus Sulfur Recovery and Tail Gas Hydrogenation followed by Recycle to the Gasifier
- CO_2 Dehydration and Pressurization (the pressurization scheme was modified to include first compression to a pressure such that the CO_2 stream forms a liquid when cooled against the cooling water, followed by pumping to the final pressure in order to reduce the parasitic load of pressurization)
- Gas Turbine (based on GE F class technology)
- Heat Recovery Steam Generator Unit
- Reheat Steam Cycle

The plant also has the necessary utilities (e.g. cooling water supply, make-up water treatment, plant and instrument air) to support the process units. Detailed process description of this case may be found in the previously referenced DOE/NETL report.

Warm Gas Cleanup

The IGCC plant employing the sorbent CO_2 capture consists of the following plant subsystems:

- Air Separation Unit (ASU)
- Coal Feed Preparation
- Gasification (based on Phillips 66 technology)
- High Temperature Syngas Cooling and Scrubbing
- Warm Gas Cleanup similar to RTI's process including Acid (H_2SO_4) Unit
- Sour Shifting
- Regenerable Sorbent CO_2 Capture (based on TDA technology)
- CO_2 Purification and Pressurization
- Gas Turbine (based on GE F class technology)
- Heat Recovery Steam Generator Unit
- Reheat Steam Cycle

In this final analysis we evaluated two cases that used 2.4 bed volumes of steam purge at two different desorption pressures (9.7 barA and 4.1 barA) along with co-current depressurization to recover and recycle the syngas trapped in the voids and pores of the sorbent back to the feed side. The overall plant efficiency for the two cases were similar however, the higher regeneration pressure case would have a lower plant cost due to savings in the compressor costs. Hence, we selected this higher pressure case for detailed cost analysis and the process description for this case is provided below.

The overall configuration of the IGCC plant with TDA's high temperature PSA-based CO_2 capture system is shown in Figure 68 while the major stream data are presented in Table 21. Plant subsystems that are different from the Cold Gas Cleanup case are described in the following.

As in the cold gas cleanup case, 6.2% of the raw syngas exiting the scrubber is compressed and recycled back to the gasifier system as quench gas.

Remainder of the scrubbed gas is preheated to a temperature of 260°C in a feed/effluent exchanger and supplied to a warm gas cleanup unit similar to RTI's process for removal of sulfur compounds utilizing a zinc titanate adsorbent in a fluidized bed. The performance of this unit as well as the production of H_2SO_4 from the SO_2 in the regenerator off-gas was developed utilizing information available in the public domain. The regenerator off-gas after particulate removal is depressurized by expansion in a power recovery turbine before feeding it to the H_2SO_4 unit. The on-site ASU provides the small amount of O_2 as required by the H_2SO_4 unit in addition to supplying oxygen to the gasifier and the catalytic combustor used for CO_2 purification (combust the residuals amounts of H_2, CO and CH_4). The hot syngas leaving the desulfurizer is cooled to a temperature of 233°C in the feed/effluent exchanger where the desulfurizer feed gas is preheated.

The clean gas is then treated in a sour shift unit similar to the Cold Gas Cleanup case consisting of two adiabatic beds in series with intercooling where intermediate pressure (IP) and medium pressure (MP) steam is generated. Steam required by the shift unit is extracted from the steam cycle.

The shifted syngas leaving the last shift reactor at a temperature of 257°C is cooled to generate MP steam, combined with recycle gas from TDA's PSA unit and then fed to the TDA's high temperature PSA unit for decarbonizing the syngas before it is combusted in the gas turbines as depicted in Figure 69. More than 97.5% of the syngas enters this decarbonizing unit where 98.5% of the CO_2 entering with the syngas is separated on a per-pass basis with the overall carbon capture being 90%. Remainder of the syngas is sent directly to the gas turbine bypassing TDA's CO2 capture system. Regeneration is accomplished utilizing steam at a desorption pressure of 9.7 barA. Two streams are regenerated, one consisting of "raw CO_2," a mixture of CO_2, steam and small amounts of residual syngas at a temperature of 186°C, and the other recycle gas," with significant amounts of other syngas components (mainly H_2) at a temperature of 228°C for recycle to the CO_2 separation unit. The raw CO_2 is cooled in a series of heat exchangers while generating low pressure (LP) steam, vacuum condensate/ makeup BFW heating and finally trim cooled against cooling water before it is compressed, preheated in a feed/effluent exchanger and then fed to a catalytic (noble metal) combustor along with O_2 to oxidize the small amounts of combustibles present in the raw CO_2 stream. The effluent from this combustor after generating HP steam is cooled in the feed/effluent exchanger. This is followed by vacuum condensate/ makeup BFW heating and finally trim cooled against cooling water. It is then further compressed to a pressure such that the CO_2 stream forms a liquid when cooled against the cooling water. This liquid CO_2 stream leaving the cooling water exchanger is then pumped to the final pressure (note that a similar scheme was used in the Cold Gas Cleanup case to maintain consistency).

The decarbonized syngas leaving the CO_2 separation (adsorption) unit at a temperature of 203°C with its accompanying unreacted steam is supplied to the gas turbines along with pressurized N_2 from the ASU. However, the amount of N_2 added to the gas turbine is significantly lower than that in the Cold Gas Cleanup case due to the large amount of water vapor present in the syngas. The combined cycle design is similar to the design in the Cold Gas Cleanup case that uses a reheat steam cycle.

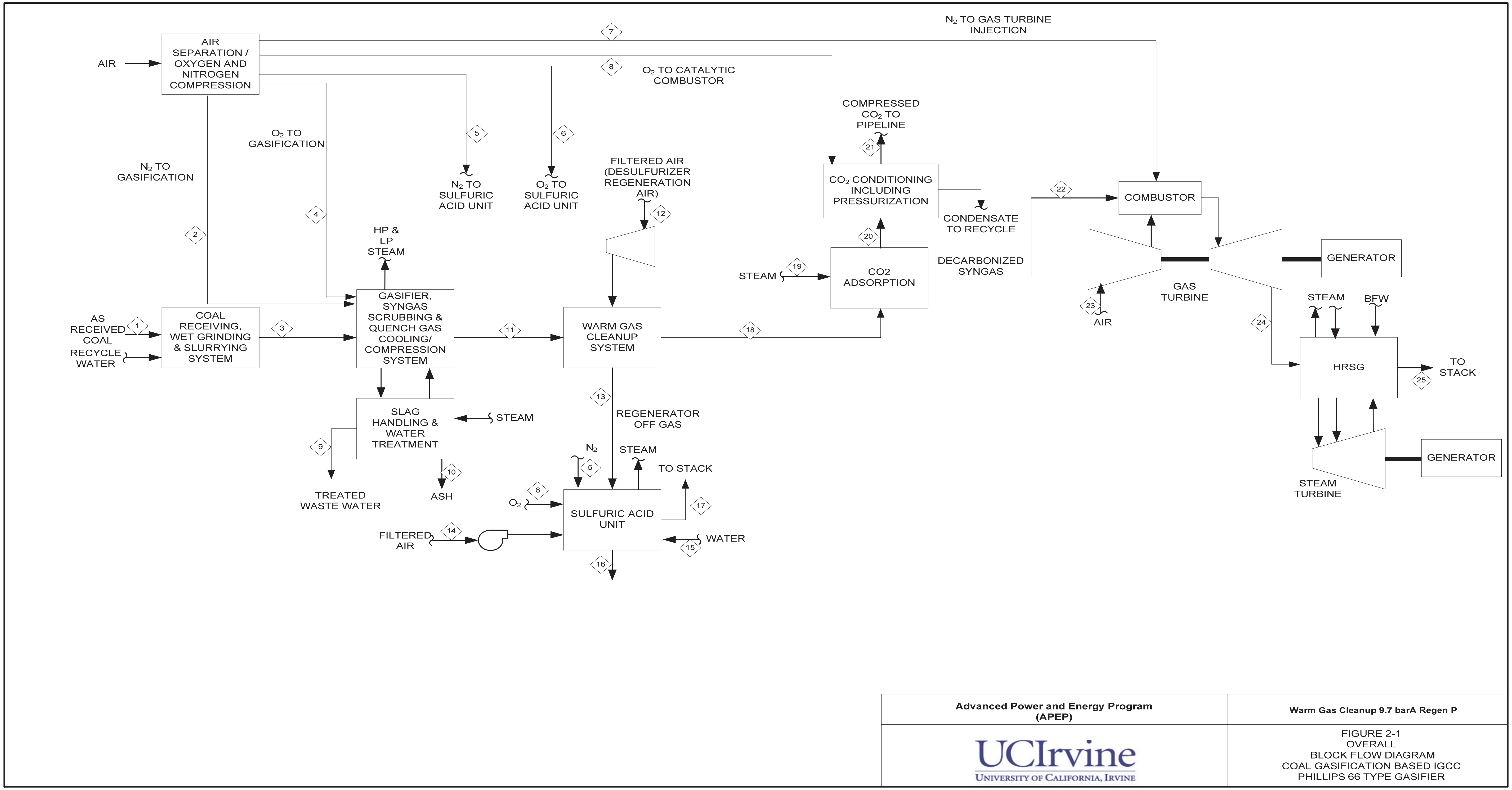


Figure 68 Block Flow Diagram for IGCC Plant with TDA's high-temperature PSA-based CO_2 Capture System

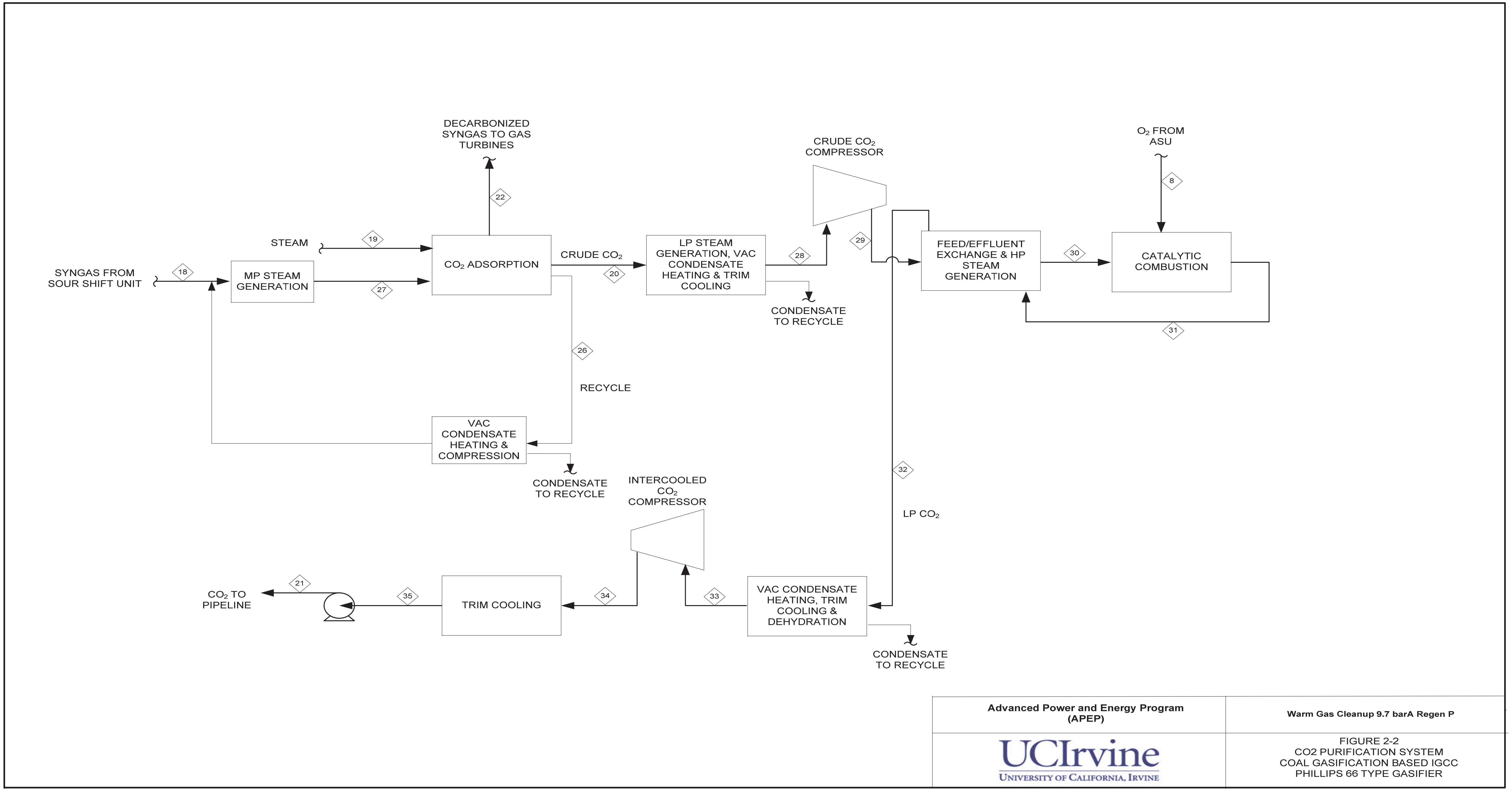


Figure 69 Block flow diagram of TDA's high temperature PSA-based CO_2 capture system along with the CO_2 purification and compression system

Table 21 Stream Data Warm Gas Cleanup 9.7 barA Regeneration Pressure

Stream number	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**
Temperature C	15		149	92	60	32	196	145	58	60	172	15	345	15	141	20	43
Pressure bar	1.01		57.92	50.33	41.2	8.62	30.52	20.5	3.1	1.01	38.78	1.01	2.07	1.01	8	1.01	1.01
Mass Flow kg/hr (Solid)	208786	0	178407	0	0	0	0	0	5305	18672	0	0	0	0	0	0	0
Vapor Frac (w/o Solids)	0		0.51	1	1	1	1	1	0.014		1	1	1	1	0	0	1
Mole Flow kmol/hr (w/o Solids)	1450	0	12668	5709	298	111	4600	171	1385		29659	1400	1594	558	178	237	2059
Mass Flow kg/hr (w/o Solids)	26122	0	152998	183743	8373	3571	129094	5492	25297		602392	40387	51484	16106	3206	18934	57223
Mole Frac (w/o Solids)																	
O2			0.04	0.95	0.005	0.95	0.005	0.95				0.208	0.011	0.208			0.034
N2			0.008	0.018	0.992	0.018	0.992	0.018			0.004	0.772	0.863	0.772			0.904
AR				0.032	0.002	0.033	0.002	0.033			0.007	0.009	0.009	0.009			0.011
H2			0.414								0.292						
CO											0.274						
CO2											0.153						0.001
H2O	1		0.538						0.983		0.238	0.011	0.002	0.01	1	0.228	0.051
CH4											0.017						
H2S + COS											0.006						
NH3 + HCN									0.003		0.008						
HCl									0.014								
SO2													0.115				
H2SO4																0.772	

Stream number	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**	**27**	**28**	**29**	**30**	**31**	**32**	**33**	**34**	**35**
Temperature C	257	205	186	39	203	15	569	154	228	198	27	98	303	426	141	27	76	27
Pressure bar	34.16	10.34	9.66	152.7	32.89	1.01	1.05	0.99	9.89	33.82	8.9	20	19.79	19.1	18.55	17.79	91	90.66
Vapor Frac	1	1	1	0	1	1	1	1	1	1	1	1	1	1	1	1	1	0
Mole Flow kmol/hr	38256	13823	21491	11144	29954	119703	146518	146518	3672	41300	11457	11457	11457	11483	11483	11142	11142	11142
Mass Flow kg/hr	750172	249024	671562	490334	316218	3453421	3898731	3898731	96867	835659	490935	490935	490934	496386	496386	490246	490246	490246
Mole Frac																		
O2						0.208	0.111	0.111										
N2	0.003				0.004	0.772	0.663	0.663	0.002	0.0032								
AR	0.005			0.001	0.006	0.009	0.009	0.009	0.004	0.0048				0.001	0.001	0.001	0.001	0.001
H2	0.404		0.013		0.507				0.299	0.4012	0.025	0.025	0.025					
CO	0.008				0.01				0.006	0.0078								
CO2	0.303		0.517	0.999	0.016		0.009	0.009	0.5	0.3251	0.97	0.97	0.97	0.969	0.969	0.999	0.999	0.999
H2O	0.26	1	0.461		0.442	0.011	0.209	0.209	0.18	0.2415	0.004	0.004	0.004	0.03	0.03			
CH4	0.012				0.016				0.009	0.0124	0.001	0.001	0.001					
NH3	0.004		0.007															

4.8.5 Results and Discussion

The plant performance of the Calibration Case (was first modeled in Aspen Plus® to compare its overall plant thermal performance with the DOE/NETL study Case 4 in order to validate the Aspen Plus® IGCC system model) is summarized in Table 22. The calculated net thermal efficiency of this Cold Gas Cleanup Case is slightly higher at 31.62% versus 31.0% (on a coal HHV basis) for the DOE/NETL study Case 4. The primary reason for the higher efficiency is that the CO_2 pressurization was accomplished by a combination of gas compression and liquid CO_2 pumping which is more efficient than just gas compression; as the process used in the previous DOE/NETL study.

As stated previously, two cases were evaluated for the Warm Gas Cleanup, each with a different regeneration pressure (9.7 barA and 4.1 barA). The performances of these two Warm Gas Cleanup Cases are summarized in Table 23 (which was developed by further modifying the Aspen Plus® IGCC system model developed for the above Cold Gas Cleanup Case). The resulting efficiencies for these cases are essentially the same at 34% (on a coal HHV basis) which is significantly higher than that for the Cold Gas Case, or an increase of as much as 7.5% in the heat rate. Low pressure regeneration has much lower steam consumption for purge however, regeneration of the CO_2 adsorber (PSA) at the lower pressure did not show any significant improvement in the overall IGCC thermal performance and the overall plant cost would be higher due to higher compressor cost for the low pressure regeneration case. The plant water consumption on a net kW generated basis is also reduced significantly for the Warm gas Cleanup Cases (11.2 vs 11.8 gpm/MWe) saving 8.5 mol of water per kWh compared to cold gas cleanup case.

It should be noted that the performance of the gas turbine for each of these cases was estimated by UCI and it is recommended that in a more detailed phase of this development program, gas turbine vendors be contacted to obtain better performance data. The effective LHV of the syngas provided to the gas turbines was held constant for each of the cases by adjusting the amount of diluent, i.e., moisture introduced into the syngas by the humidification operation in the case of cold gas cleanup and amount of N_2 (supplied by the ASU) in the case of the warm gas cleanup. The molar ratio of moisture to N_2 is significantly higher for the warm gas cleanup case and since H_2O has a much higher specific heat than N_2, it may be expected that the NO_x emission for this case will be significantly lower.

We also carried out comparison of the warm gas cleanup case with high pressure regeneration against the cold gas cleanup case on the basis of same coal feed rate of 216,676.5 kg/h and the performance results are summarized in Table 24 and Table 25 respectively. Cold gas capture case produced higher total power than warm gas capture. This is due to more power being generated in the gas turbine for the cold gas cleanup case. Water vapor is the major diluent present in the syngas entering gas turbine in the warm gas cleanup case while N_2 from ASU is the major diluent present in the syngas entering the gas turbine in the cold gas cleanup case. Less amount of water vapor is required to achieve the same level of NOx emissions as with N_2. This reduces the gas turbine output for the warm gas cleanup case when expressed on a unit coal flow basis.

However, the net power produced is higher for the warm gas cleanup with lower auxiliary loads (121,004 kWe) in the case of warm gas cleanup with TDA's CO_2 capture system compared to cold gas cleanup (176,392 kWe) with Selexol^{TM} system providing a net difference in auxiliary load of 55.4 MWe. There are two major subsystems that result in this reduction in auxiliary load for the warm gas cleanup case the ASU and the CO_2 capture system.

In the case of the ASU the auxiliary load for cold gas cleanup case is very high (109.3 MWe) compared to warm gas cleanup case (70.7 MWe) (a net difference of 38.6 MWe). The ASU was modeled as comprising a low pressure (LP) ASU and an elevated pressure (EP) ASU for both the cases. The relative size of these two units is set by the amount of N_2 required for gas turbine injection to minimize compression of air. The warm gas cleanup case requires less N_2 for the gas turbine and thus the size of the EP ASU unit is smaller (while that of the LP ASU is larger) when compared to cold gas cleanup case. Hence the auxiliary load for air compression is higher for cold gas cleanup case. Also the diluent N_2 supplied by the ASU to the gas turbine requires a significant amount of compression (from pressure leaving the elevated pressure cold box to that required by the gas turbine). Hence the auxiliary load for the nitrogen compressor is higher for cold gas capture. However, the auxiliary load for oxygen compressor is higher for warm gas cleanup case since more oxygen from the LP ASU needs to be compressed for use in the gasifier and the catalytic combustor.

The auxiliary load for CO_2 removal, purification and compression for the warm gas cleanup case with TDA's high temperature PSA system is significantly lower at 25.4 MWe compared to 39.5 MWe for cold gas cleanup with Selexol^{TM} (a net difference of 14.1 MWe). When including the auxiliary loads for sulfur removal subsystems the difference between warm gas cleanup and cold gas cleanup is even higher at 16.3 MWe.

The above results indicate that warm gas cleanup technology coupled with the CO_2 adsorption process being developed by TDA can make a substantial improvement in the IGCC plant thermal performance designed for approaching near zero emissions, i.e., to include CO_2 capture.

The plant cost and the levelized cost of electricity estimates for the Cold Gas Cleanup Case and the Warm gas Cleanup Case with the CO_2 adsorber (PSA) regeneration performed at the higher pressure of 9.7 barA are presented in Table 26 through Table 29. The plant cost estimate for the Cold Gas Cleanup Case is lower than that for the DOE/NETL study Case 4. The primary reason again for the lower cost is that the savings in the CO_2 pressurization cost which was accomplished, as mentioned above, by a combination of gas compression and liquid CO_2 pumping, liquid pumps being significantly cheaper than gas compressors. The resulting plant cost for the Warm gas Cleanup Case is $2,418/kW while that for the Cold Gas Case is $2,754/kW, or a decrease of as much as 12% over the Cold Gas Cleanup Case. The levelized cost of electricity with transport, storage and monitoring (TS&M) costs for the CO_2 included is $92.9/MWh for the Warm gas Cleanup Case while that for the Cold Gas Case is $105.2/MWh, or a decrease of also 12% over the Cold Gas Cleanup Case.

The avoided cost of capturing the CO_2 calculated for both the cold gas cleanup case and the warm gas cleanup case at higher regeneration pressure of 9.7 barA are presented in Table 30. As can be seen, again, the warm gas cleanup with TDA's PSA technology can capture CO_2 at a cost of $31.12 per tonne CO_2 avoided compared to $49.50 per tonne CO_2 avoided for cold gas cleanup with Selexol^{TM} technology. Thus TDA's CO_2 capture technology captures CO_2 at a 37.1% lower cost than Selexol^{TM} based technology.

Table 22 Overall Plant Performance Summary, Cold Gas Cleanup Case-Fully Loaded GTs

	UNITS	
GAS TURBINE POWER	kWe	464,000
STEAM TURBINE POWER	kWe	227,247
TOTAL POWER	kWe	691,247
TOTAL AUXILIARY CONSUMPTION	kWe	175,994
NET POWER OUTPUT	kWe	515,253
AUXILIARY LOAD SUMMARY		
COAL HANDLING	kWe	439
COAL MILLING	kWe	2,225
COAL SLURRY PUMPS	kWe	581
SLAG HANDLING & DEWATERING	kWe	1,159
AIR SEPARATION UNIT AUXILIARIES	kWe	1,039
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	kWe	63,739
OXYGEN COMPRESSOR	kWe	8,873
NITROGEN COMPRESSOR	kWe	34,284
SYNGAS RECYCLE COMPRESSOR	kWe	1,117
TAIL GAS RECYCLE COMPRESSOR	kWe	3,190
CO2 COMPRESSOR	kWe	19,849
BOILER FEEDWATER & DEMIN PUMPS	kWe	5,723
VACUUM CONDENSATE PUMP	kWe	353
PROCESS CONDENSATE & SWS SYSTEMS	kWe	299
HUMIDIFIER & BFW CIRCULATING PUMPS	kWe	44
COOLING WATER CIRCULATING PUMPS	kWe	4,286
COOLING TOWER FANS	kWe	2,290
SCRUBBER PUMPS	kWe	70
SELEXOL UNIT	kWe	19,582
GAS TURBINE AUXILIARIES	kWe	1,000
STEAM TURBINE AUXILIARIES	kWe	99
CLAUS & TAIL GAS TREATING AUXILIARIES	kWe	199
MISCELLANEOUS BALANCE OF PLANT	kWe	2,993
TRANSFORMER LOSSES	kWe	2,560
NET PLANT EFFICIENCY, HHV	%	31.62
NET HEAT RATE	kJ/kWH	11,387
	BTU/kWH	10,793
CONDENSER COOLING DUTY	10^6 kJ/H	1,254
	10^6 BTU/H	1,189
CONSUMABLES		
AS-RECEIVED COAL FEED	kg/H	216,187
	LB/H	476,692
THERMAL INPUT	kWt HHV	1,629,385
RAW WATER USAGE	M^3/MIN	23.07
	GPM	6,094
CARBON CAPTURED	%	90

Table 23 Overall Plant Performance Summary Warm Gas Cleanup Cases- Fully Loaded GTs

CASE DESIGNATION	**UNITS**	**Regeneration Pressure**	
		9.7 barA	**4.1 barA**
GAS TURBINE POWER	kWe	464,000	464,000
STEAM TURBINE POWER	kWe	269,028	285,417
TOTAL GROSS POWER	kWe	733,028	749,417
TOTAL AUXILIARY CONSUMPTION	kWe	131,163	147,894
NET POWER OUTPUT	kWe	601,865	601,523
AUXILIARY LOAD SUMMARY			
COAL HANDLING	kWe	477	475
COAL MILLING	kWe	2,417	2,409
COAL SLURRY PUMPS	kWe	631	629
SLAG HANDLING & DEWATERING	kWe	1,200	1,196
AIR SEPARATION UNIT AUXILIARIES	kWe	258	370
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	kWe	45,678	48,898
OXYGEN COMPRESSOR	kWe	20,091	18,702
NITROGEN COMPRESSOR	kWe	9,400	13,431
SYNGAS RECYCLE COMPRESSOR	kWe	1,210	1,207
CO2 PURIFICATION & COMPRESSION	kWe	27,572	38,074
BOILER FEEDWATER & DEMIN PUMPS	kWe	6,504	6,405
VACUUM CONDENSATE PUMP	kWe	442	412
PROCESS CONDENSATE & SWS SYSTEMS	kWe	8	7
BFW CIRCULATING PUMPS	kWe	104	103
COOLING WATER CIRCULATING PUMPS	kWe	4,380	4,608
COOLING TOWER FANS	kWe	2,340	2,462
SCRUBBER PUMPS	kWe	76	76
DESULFURIZER UNIT	kWe	5,304	5,285
GAS TURBINE AUXILIARIES	kWe	1,000	1,000
STEAM TURBINE AUXILIARIES	kWe	117	124
H2SO4 UNIT	kWe	-4,013	-3,997
MISCELLANEOUS BALANCE OF PLANT	kWe	3,252	3,241
TRANSFORMER LOSSES	kWe	2,715	2,776
NET PLANT EFFICIENCY, HHV	%	33.99	34.08
NET HEAT RATE	kJ/KWH	10,591	10,563
	BTU/kWH	10,038	10,012
CONDENSER COOLING DUTYNET POWER	10^6 kJ/H	1,449	1,514
	10^6 BTU/	1,373	1,435
CONSUMABLES			
AS-RECEIVED COAL FEED	kg/H	234867	234112
	LB/H	517882	516218
THERMAL INPUT	kWt HHV	1770177	1764488
RAW WATER USAGE	M^3/MIN	25.41	24.76
	GPM	6,714	6,543
CARBON CAPTURED	%	90	90

Table 24 Overall Plant Performance Summary Cold Gas Cleanup Case -Coal Feed Rate=216,676.5 kg/hr

	UNITS	
GAS TURBINE POWER	kWe	465,051
STEAM TURBINE POWER	kWe	227,761
TOTAL POWER	kWe	692,812
TOTAL AUXILIARY CONSUMPTION	kWe	176,392
NET POWER OUTPUT	kWe	516,420
AUXILIARY LOAD SUMMARY		
COAL HANDLING	kWe	440
COAL MILLING	kWe	2230
COAL SLURRY PUMPS	kWe	582
SLAG HANDLING & DEWATERING	kWe	1162
AIR SEPARATION UNIT AUXILIARIES	kWe	1042
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	kWe	63,883
OXYGEN COMPRESSOR	kWe	8,893
NITROGEN COMPRESSOR	kWe	34,362
SYNGAS RECYCLE COMPRESSOR	kWe	1,119
TAIL GAS RECYCLE COMPRESSOR	kWe	3,197
CO2 COMPRESSOR	kWe	19,894
BOILER FEEDWATER & DEMIN PUMPS	kWe	5,735
VACUUM CONDENSATE PUMP	kWe	354
PROCESS CONDENSATE & SWS SYSTEMS	kWe	300
HUMIDIFIER & BFW CIRCULATING PUMPS	kWe	44
COOLING WATER CIRCULATING PUMPS	kWe	4,296
COOLING TOWER FANS	kWe	2,295
SCRUBBER PUMPS	kWe	70
SELEXOL UNIT	kWe	19,627
GAS TURBINE AUXILIARIES	kWe	1,002
STEAM TURBINE AUXILIARIES	kWe	99
CLAUS & TAIL GAS TREATING AUXILIARIES	kWe	199
MISCELLANEOUS BALANCE OF PLANT	kWe	3,000
TRANSFORMER LOSSES	kWe	2,566
NET PLANT EFFICIENCY, HHV	%	31.62
NET HEAT RATE	kJ/kWH	11,387
	BTU/kWH	10,793
CONDENSER COOLING DUTY	10^6 kJ/H	1,257
	10^6 BTU/H	1,191
CONSUMABLES		
AS-RECEIVED COAL FEED	kg/H	216,676.50
	LB/H	477,772
THERMAL INPUT	kWt HHV	1,633,075
RAW WATER USAGE	M^3/MIN	23.12
	GPM	6,108
CARBON CAPTURED	%	90

Table 25 Overall Plant Performance Summary Warm Gas Cleanup Case -Coal feed rate of 216,676.5 kg/h

Regeneration Pressure (9.7 barA)	**UNITS**	
GAS TURBINE POWER	kWe	428,063
STEAM TURBINE POWER	kWe	248,191
TOTAL POWER	kWe	676,254
TOTAL AUXILIARY CONSUMPTION	kWe	121,004
NET POWER OUTPUT	kWe	555,250
AUXILIARY LOAD SUMMARY		
COAL HANDLING	kWe	440
COAL MILLING	kWe	2230
COAL SLURRY PUMPS	kWe	582
SLAG HANDLING & DEWATERING	kWe	1107
AIR SEPARATION UNIT AUXILIARIES	kWe	238
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	kWe	42,140
OXYGEN COMPRESSOR	kWe	18,535
NITROGEN COMPRESSOR	kWe	8.672
SYNGAS RECYCLE COMPRESSOR	kWe	1,117
CO2 PURIFICATION & COMPRESSION	kWe	25,437
BOILER FEEDWATER & DEMIN PUMPS	kWe	6,000
VACUUM CONDENSATE PUMP	kWe	408
PROCESS CONDENSATE & SWS SYSTEMS	kWe	7.79
BFW CIRCULATING PUMPS	kWe	95.5
COOLING WATER CIRCULATING PUMPS	kWe	4,040
COOLING TOWER FANS	kWe	2,158
SCRUBBER PUMPS	kWe	70
DESULFURIZER UNIT	kWe	4,893
GAS TURBINE AUXILIARIES	kWe	923
STEAM TURBINE AUXILIARIES	kWe	108
H2SO4 UNIT	kWe	-3,702
MISCELLANEOUS BALANCE OF PLANT	kWe	3,000
TRANSFORMER LOSSES	kWe	2,505
NET PLANT EFFICIENCY, HHV	%	33.99
NET HEAT RATE	kJ/kWH	10,591
	BTU/kWH	10,038
CONDENSER COOLING DUTY	10^6 kJ/H	1,337
	10^6 BTU/H	1267
CONSUMABLES		
AS-RECEIVED COAL FEED	kg/H	216,676.50
	LB/H	477,772
THERMAL INPUT	kWt HHV	1,633,075
RAW WATER USAGE	M^3/MIN	23.45
	GPM	6,194
CARBON CAPTURED	%	90

Table 26 Overall Plant Cost Summary Cold Gas Cleanup Case-Fully Loaded GTs

UNIT	2007 Installed Cost ($1000)
ASU	220,067
Fuel receiving, preparation & feeding	92,151
Gasifier, syngas cooler & aux	248,837
Gasification foundations	18,067
Ash handling systems	37,123
Flare stack system	3,008
Shift reactor	13,534
Syngas scrubber (included in gasification auxiliaries)	0
Blowback gas systems	1,330
Fuel gas piping	1,574
Gas cleanup foundations	1,703
Hg Removal + LTGC	46,712
Selexol	199,729
Claus + TG Recycle	33,288
CO2 compression, dehydration + pumping	24,586
Gas turbine + generator + auxiliaries	132,015
HRSG, ducting + stack	56,119
Steam turbine + generator + auxiliaries	50,719
Surface condenser	6,984
Feedwater system	19,616
Water makeup + pretreating	2,035
Other feedwater subsystems	3,208
Service water systems	5,842
Other boiler plant systems	6,468
Fuel oil system & nat gas	1,915
Waste water treatment	2,251
Misc. power plant equipment	2,537
Cooling water system	35,313
Accessory electric plant	87,698
Instrumentationo & controls	27,099
Improvement to site	19,578
Buildings & structures	17,936
Total	1,419,040
Total $/kW	**2,754**

Table 27 Overall Plant Cost Summary
Warm Gas Cleanup Case - 9.7 barA Regeneration Pressure – Fully Loaded GTs

UNIT	**2007 Installed Cost ($1000)**
ASU	243,864
Fuel receiving, preparation & feeding	97,655
Gasifier, syngas cooler & aux	263,700
Gasification foundations	19,146
Ash handling systems	39,340
Flare stack system	3,187
Warm gas desulfurization	26,610
H2SO4 unit	58,139
Shift reactor	13,683
Syngas scrubber (included in gasification auxiliaries)	0
Blowback gas systems	1,410
Fuel gas piping	2,381
Gas cleanup foundations	1,721
Hg Removal, CO2 separation / Recycle	104,507
CO2 purification / heat recovery	27,887
CO2 compression / drying / pumping	51,076
Gas turbine + generator + auxiliaries	132,015
HRSG, ducting + stack	58,145
Steam turbine + generator + auxiliaries	57,080
Surface condenser	7,727
Feedwater system	24,265
Water makeup + pretreating	2,178
Other feedwater subsystems	3,610
Service water systems	6,191
Other boiler plant systems	7,279
Fuel oil system & nat gas	1,915
Waste water treatment	2,385
Misc. power plant equipment	2,643
Cooling water system	35,848
Accessory electric plant	91,376
Instrumentationo & controls	28,718
Improvement to site	20,747
Buildings & structures	19,008
Total	1,455,436
Total $/kW	**2,418**

Table 28 Cost of Electricity Cold Gas Cleanup Case – Fully Loaded GTs 2007 $

Net power, MW	515.25
Capacity factor (CF), %	80
Total plant cost (TPC), $	1,419,040,277
6 month labor cost	12,883,919
1 month maintenance materials	2,213,213
1 month non-fuel consumables	531,751
1 month waste disposal	290,027
25% of 1 month fuel cost at 100% CF	1,660,760
2% of TPC	28,380,806
60 day supply of fuel & consumables at 100% CF	14,153,013
0.5% of TPC (spare parts)	7,095,201
Initial catalyst & chemicals cost, $	7,180,117
Land	900,000
Other owners's costs (15% of TPC)	212,856,041
Financing costs	38,314,087
Total overnight cost (TOC), $	1,745,499,213
Fixed operating cost for initial year of operation (OCF), $	54,148,643
Annual feed cost at above CF for initial year (OCV1), $	63,773,186
Other annual variable operating cost at above CF for initial year (OCV2), $	29,135,919
Annual CO2 transporting, storing, and monitoring cost at above CF for initial year (OCV3), $	19,419,868
Annual byproduct revenues at above CF for initial year (OCV4), $	3,749,093
1st year cost of electricity (COE) w/o CO2 TS&M, $/MWh	**99.8**
1st year cost of electricity (COE), $/MWh	**105.2**

Table 29 Cost of Electricity Warm Gas Cleanup Case - 9.7 barA Regeneration Pressure – Fully Loaded GTs 2007 $

Net power, MW	601.87
Capacity factor (CF), %	80
Total plant cost (TPC), $	1,455,435,703
6 month labor cost	12,850,781
1 month maintenance materials	2,205,008
1 month non-fuel consumables	1,075,553
1 month waste disposal	393,066
25% of 1 month fuel cost at 100% CF	1,804,263
2% of TPC	29,108,714
60 day supply of fuel & consumables at 100% CF	16,358,015
0.5% of TPC (spare parts)	7,277,179
Initial catalyst & chemicals cost, $	19,684,704
Land	900,000
Other owners's costs	218,315,355
Financing costs	39,296,764
Total overnight cost (TOC), $	1,804,705,104
Fixed operating cost for initial year of operation (OCF), $	54,810,276
Annual feed cost at above CF for initial year (OCV1), $	69,283,698
Other annual variable operating cost at above CF for initial year (OCV2), $	35,266,816
Annual CO2 transporting, storing, and monitoring cost at above CF for initial year (OCV3), $	21,139,715
Annual byproduct revenues at above CF for initial year (OCV4), $	13,166,137
1st year cost of electricity (COE) w/o CO2 TS&M, $/MWh	**87.8**
1st year cost of electricity (COE), $/MWh	**92.9**

Table 30 Avoided CO2 Costs (2007 $)

Gasifier	**Phillips 66**	
Syngas Cleanup	**Cold Gas**	**Warm Gas**
Decarbonization Technology	**Selexol**	**TDA PSA**
Net Power, MW	515.25	601.87
CO2 Emitted, ST/h	552.528	600.27
CO2 Emitted, ST/MWh	1.072	0.997
COE, $/MWh	105.15	92.9
Reference IGCC Case		
Avoided Cost (1st Year), $/ST	44.9	28.23
Avoided Cost (1st Year), $/tonne	49.5	31.12
Reference SCPC Case		
Avoided Cost (1st Year), $/ST	60.82	44.62
Avoided Cost (1st Year), $/tonne	67.05	49.18
Reference IGCC w/o CO2 capture		
Net Power, MW	625.06	
CO2 Emitted, ST/MWh	0.855	
COE, $/MWh (See Notes 1 & 2)	71.58	
Reference SCPC w/o CO2 capture		
Net Power, MW	549.99	
CO2 Emitted, ST/MWh	0.884	
COE, $/MWh (See Note 1)	57.91	

Notes

1. Variable cost corrected for capacity factor
2. Byproduct sulfur credit taken since PSA cases take credit for H_2SO_4 produced.

4.8.6 Summary of Process Design and System Analysis

The Warm Gas Cleanup system using TDA's high temperature PSA-based CO_2 capture system achieved a net plant efficiency of 34% (on a coal HHV basis). This net plant efficiency at 90% carbon capture is significantly higher than that can be achieved for the Cold Gas Case using Selexol^{TM} scrubber at 31.6%, corresponding to a 7.5% decrease in the heat rate for TDA's Warm Gas Cleanup system. The plant water consumption on a net kW generated basis is also reduced significantly for TDA's Warm gas Cleanup process (a reduction of 8.5 kmol per MWh), preserving a valuable resource.

The capital expense for the plant was estimated following the "Cost Guidelines provide by DOE/NETL (2007) (e.g. higher contingencies were applied for unproven technologies such as the high temperature PSA unit.) The plant cost for TDA's Warm gas Cleanup Process is estimated as $2,418/kW, which is 12% lower than that of the Cold Gas Cleanup Case at $2,754/kW. The levelized cost of electricity including the transport, storage and monitoring (TS&M) costs for CO_2 is calculated as $92.9/MWh for TDA's Warm gas Cleanup Case. This cost for the Cold Gas Case using the Selexol^{TM} scrubbing technology is calculated as $105.2/MWh.

Table 31 Comparison of TDA's Warm Gas Cleanup against SelexolTM based Cold Gas Cleanup System. Basis: IGCC plant operating with fully loaded GE F class gas turbine that generates 464 MWe power

	Cold Gas Cleanup Selexol^{TM}	**Warm Gas Cleanup TDA's CO_2 Sorbent**
CO_2 Capture, %	90.0	90.0
Gross Power Generated, kWe	691,247	733,028
Gas Turbine Power	464,000	464,000
Steam Turbine Power	227,247	269,028
Auxiliary Load, kWe	175,994	131,163
Net Power, kWe	515,253	601,865
Net Plant Efficiency, % HHV	**31.6%**	**34.0%**
Coal Feed Rate, kg/h	216,187	234,867
Raw Water Usage, GPM/MWe	11.8	11.2
Total Plant Cost, $/kWe	2,754	2,418
COE without CO_2 TS&M, $/MWh	99.8	87.8
COE with CO2 TS&M, $/MWh	**105.2**	**92.9**

The results of the system analysis in Table 31 suggest that TDA's high temperature PSA-based Warm Gas Clean-up Technology can make a substantial improvement in the IGCC plant thermal performance for achieving near zero emissions (including greater than 90% carbon capture). The capital expenses are also expected to be lower than that of Selexol's™. The higher net plant efficiency and lower capital and operating costs results in substantial reduction in the cost of electricity for the IGCC plant equipped with TDA's high temperature PSA-based carbon capture system.

4.9 Task 9. Fabrication of Prototype Test Unit

Both units were built and assembled at TDA. The vessels used for housing the sorbent were sent out to be fabricated by an ASME certified welder. The frame, tubing and control box were all assembled outside the enclosure due to a lack of space. After assembly, the apparatus was moved inside the NEMA enclosures.

Figure 70. CO_2 Adsorption Unit (Left) Gas Conditioning Unit (Right) before installing inside the NEMA Enclosures.

At the top and bottom of the sorbent vessels, heated valve boxes are located which contain a series of pneumatically actuated valves and tubing manifolds (Figure 71). The valve bodies reside inside a metal box which is heated using a flat plate heater. The valve bonnets remain outside the box due to their lower temperature rating. All valves are independently controlled via pneumatic lines. The valves are supplied by High Pressure Equipment Company and are rated to 9000 psig at 260°C.

Syngas flow as well as nitrogen for regeneration is controlled by a Badger Research Control Valve. This proportional valve is pneumatically operated and allows for very precise flow. A PID loop controls the valve based on the flow meter signal that is installed after it.

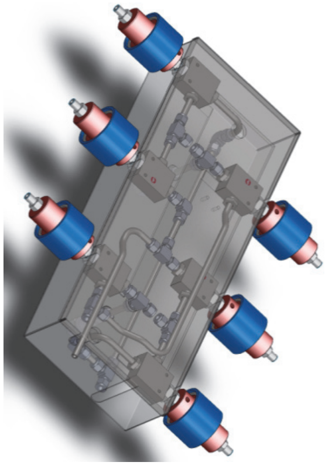


Figure 71. Valve box layout.

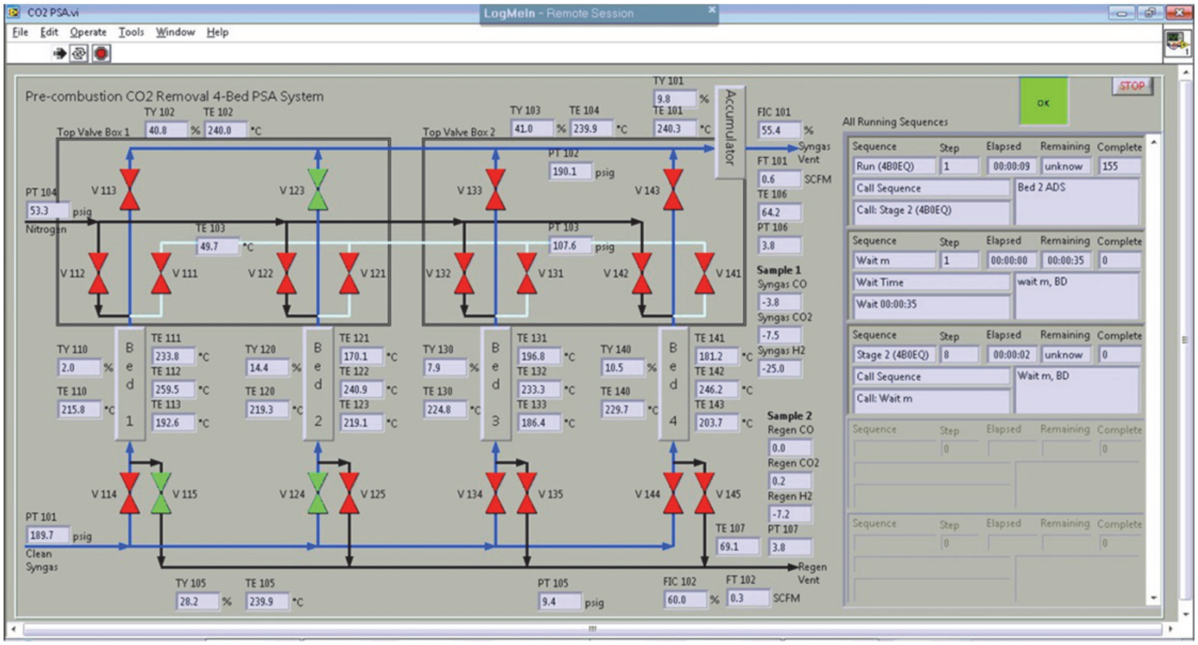
Once the units were assembled, they were then placed inside the NEMA enclosures (Figure 72) and temporary installed in a laboratory at the TDA facilities for shakedown and initial testing.

The test unit is controlled by National Instruments™ software and hardware. All I/O is handled by Compact Fieldpoint® (Figure 73). Compact Fieldpoint® is a rugged industrial programmable automation controller which is capable of withstanding the elevated temperatures that could be reached inside the enclosure. The system is controlled via a single laptop PC running LabVIEW® developmental software.

Figure 72. CO_2 Adsorption Unit (Left) Gas Conditioning Unit (Right) inside NEMA enclosures.

The LabVIEW® programming and user interface was programmed by TDA engineers. The GUI allows for flexible programming by the operators. The main front page (Figure 74) displays all critical information and the current state of the operations which are being performed by the automated software. Other pages in the GUI display real time graphs which are used to identify trends and diagnose problems. Autonomous operation is achieved by deployment of sequencers which can operate multiple simultaneous tasks. Each sequence can be

Figure 73 Compact Field Point control modules.



Pre-combustion CO2 Removal 4-Bed PSA System

TY 101 9.8 %
TE 101 240.3 °C
TY 102 40.8 %
TE 102 240.0 °C
TY 103 41.0 %
TE 104 239.9 °C
PT 102 190.1 psig
PT 103 107.6 psig
PT 104 53.3 psig
TE 103 49.7 °C
FIC 101 55.4 %
FT 101 0.6 SCFM
TE 106 64.2
PT 106 3.8

Sample 1
Syngas CO -3.8
Syngas CO2 -7.5
Syngas H2 -25.0

TY 110 2.0 %
TE 110 215.8 °C
TE 111 233.8 °C
TE 112 259.5 °C
TE 113 192.6 °C
TY 120 14.4 %
TE 120 219.3 °C
TE 121 170.1 °C
TE 122 240.9 °C
TE 123 219.1 °C
TY 130 7.9 %
TE 130 224.8 °C
TE 131 196.8 °C
TE 132 233.3 °C
TE 133 186.4 °C
TY 140 10.5 %
TE 140 229.7 °C
TE 141 181.2 °C
TE 142 246.2 °C
TE 143 203.7 °C

Sample 2
Regen CO 0.0
Regen CO2 0.2
Regen H2 -7.2
TE 107 69.1
PT 107 3.8

PT 101 189.7 psig
TY 105 28.2 %
TE 105 239.9 °C
PT 105 9.4 psig
FIC 102 60.0 %
FT 102 0.3 SCFM

All Running Sequences

Sequence	Step	Elapsed	Remaining	Complete
Run (4B0EQ)	1	00:00:09	unknow	155
Call Sequence Call: Stage 2 (4B0EQ)	Bed 2 ADS			

Sequence	Step	Elapsed	Remaining	Complete
Wait m	1	00:00:00	00:00:35	0
Wait Time Wait 00:00:35	wait m, BD			

Sequence	Step	Elapsed	Remaining	Complete
Stage 2 (4B0EQ)	8	00:00:02	unknow	0
Call Sequence Call: Wait m	Wait m, BD			

Sequence	Step	Elapsed	Remaining	Complete
	0			0

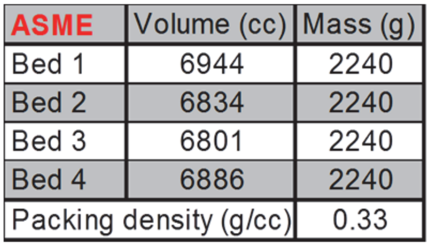
Sequence	Step	Elapsed	Remaining	Complete
	0			0

Figure 74 Control Software GUI.

individually programmed to tailor the operation according to the various operating conditions encountered in the field.

4.9.1 Tests at TDA

Once installed at TDA, the Test Unit underwent initial shakedown, which included ensuring all valves were operational, all thermocouples were reading properly, all heaters were operational and the PID loops were tuned. Each sorbent bed was filled with 2240 g of TDA's carbon sorbent. Denstone® ceramic beads were placed at the top and bottom of each vessel to contain the sorbent in the cylindrical portion of the vessel. Screened VCR gaskets were used to prevent the media from escaping the beds. Figure 75 shows the packing density of each bed in the system.



ASME	Volume (cc)	Mass (g)
Bed 1	6944	2240
Bed 2	6834	2240
Bed 3	6801	2240
Bed 4	6886	2240
Packing density (g/cc)		0.33

Figure 75. Sorbent bed loading.

4.9.2 Bed Capacity Tests

Baseline testing at TDA was performed by metering in CO_2 and N_2 into the Test Unit to simulate the partial pressure of CO_2 in the synthesis gas. The first tests performed were to simulate the partial pressure of CO_2 in NCCC's syngas. These tests were used as our baseline for CO_2 capacity of the sorbent beds.

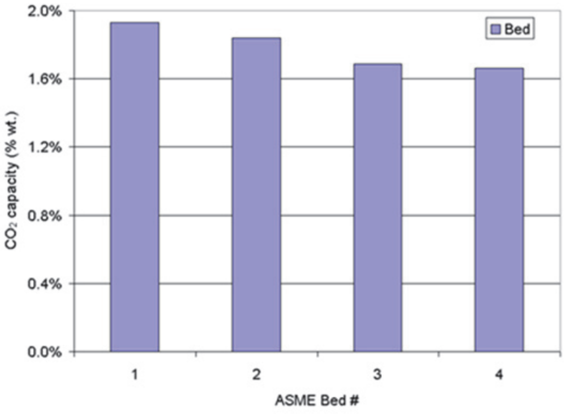


Figure 76. Single Bed Capacities before NCCC Testing. Bed Temp= 200°C, P_{ads} = 190 PSIG, P_{CO2}= 28.5 PSIG

A series of single bed cycles were conducted in order to determine the CO_2 adsorption capacity for each bed. Bottled CO_2 and N_2 were metered into the system using a pair of mass flow controllers. A CO_2 partial pressure of 28 psig was maintained throughout the testing while the sorbent beds were maintained at 190 psig during adsorption along with a temperature of 200°C. As shown in the results in Figure 76, all four beds had a CO_2 capacity between 1.66% wt. and 1.94% wt. (g of CO_2 removed per g of sorbent). The weight capacities were much lower than observed in the bench-scale experiments lower due to the low partial pressure of the CO_2 in the feed stream. These numbers will be used as the baseline performance for the NCCC testing.

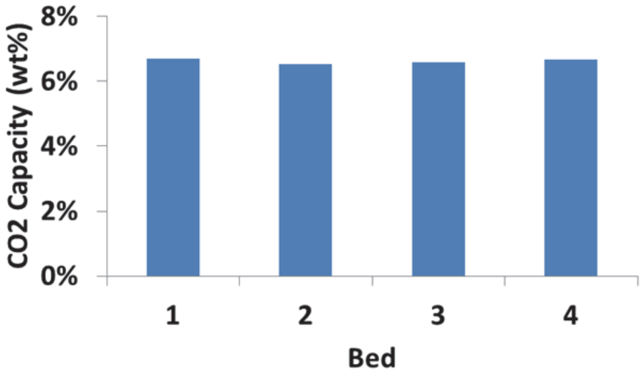


Figure 77 Laboratory testing of CO_2 Capacity for Wabash Conditions (CO_2/N_2 Single Bed Capacities), P_{CO2}=115 psig, Bed Temp = 220°C.

Once the NCCC testing was accomplished, the Test Unit was shipped back to TDA and was reinstalled in the lab to determine if the sorbent's performance has permanently been changed. These results are provided in the Field Test section of this report. Baseline tests were then performed using the Wabash gas stream conditions. Wabash's gasifier has a much higher CO content and operates at a high pressure than NCCC's. Single bed cycles were carried out to determine the CO_2 capacity of individual beds (some differences were expected due to differences in the loading process and as a result of using different sorbent batches with slightly different CO_2 capacity). CO_2 and N_2 were used to simulate the syngas and each bed was individually allowed to reach saturation and the total CO_2 weight capacity was calculated (Figure 77). The average capacity of the sorbent beds was 6.6% wt.

The WGS reactor was also tested during this time and due to the constant generation of steam, single bed tests were unable to be performed. The 4-bed cycling sequence was used to determine the weight capacity of the sorbent before shipping the unit to Wabash. CO and CO_2 and H_2 were metered into the system using mass flow controllers. The percentages of each gas were fixed but the total flow rate was allowed to fluctuate as the beds were being cycled. Steam was added in order to achieve proper WGS. The results of this baseline test are show in Figure 78. This baseline test is more indicative of the actual conditions at Wabash and will be used as the comparison numbers to chart possible decrease in performance.

The average capacity of 4% wt. is much lower than the single bed capacity of 6.6% wt. The single bed testing takes advantage of the entire sorbent bed whereas in the 4-bed cycling capacity test the beds change position well before CO_2 breakthrough to account for the two pressure equalizations hence, is unable to utilize the entire sorbent bed.

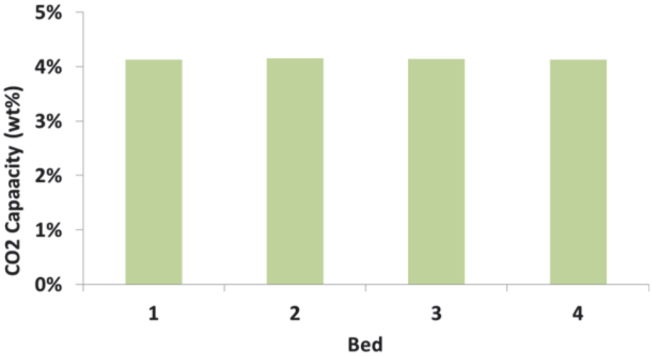


Figure 78 Laboratory testing of CO_2 Capacities for Wabash Conditions (CO_2, CO, H_2, H_2O, 4-Bed cycling), P_{CO2}=140 psig, Bed Temp = 220°C.

4.10 Task 10. Field Tests

4.10.1 Testing at the National Carbon Capture Center

The test unit was shipped to the National Carbon Capture Center (NCCC) in Wilsonville, Alabama in mid-October of 2011. The unit was installed (Figure 79) by onsite personnel along with engineers from TDA. The synthesis gas supply to the unit was taken from the exit of NCCC's gas conditioning unit (GCU). The GCU removes sulfur compounds and performs water gas shift reaction. Once installed, NCCC's gasifier underwent an unexpected shut down which delayed

Figure 79. Test Unit Installed At the NCCC.

the start of the testing period until the end of November 2011.

TDA's testing began November 22 and ending on December 7, 2011. Before NCCC's scheduled shutdown of their gasifier, the sorbent underwent 1030 cycles of adsorption and regeneration.

4.10.2 NCCC Testing Results

Figure 80 shows the CO_2 in the syngas product stream compared to the CO_2 in the incoming syngas during the initial startup of the testing. Within 2 hours the CO_2 levels dropped, leaving only small cyclic CO_2 breakthrough (we attributed it to a small leakage from one of the sorbent beds, which was driven by incomplete regeneration due to failure in one of the valves). Nevertheless, once the system reached steady state, the average CO_2 removal was over 98% with an average CO_2 capture of over 99% for the duration of the testing period.

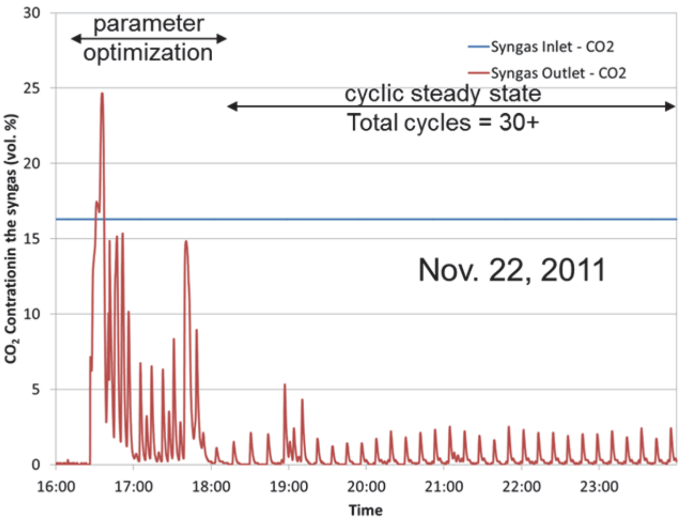


Figure 80. Raw test data showing the CO_2 concentration on volumetric basis at the outlet stream from the test unit (inlet CO_2 concentration was at 16% vol. as indicated by the blue line).

Figure 81 shows the CO_2 outlet concentration from the test unit both for the CO_2-free synthesis gas (red) and for the regeneration off-gas (blue) for approximately 82 cycles. The overall CO_2 capture was greater than 98% for this portion of the testing. The CO_2 in the synthesis gas was allowed to breakthrough to 2% to ensure the entire bed was being utilized. The CO_2 concentration in the regeneration stream peaked out at 24% and was allowed to regenerate until the CO_2 reached below 2% to ensure that the bed had sufficiently regenerated.

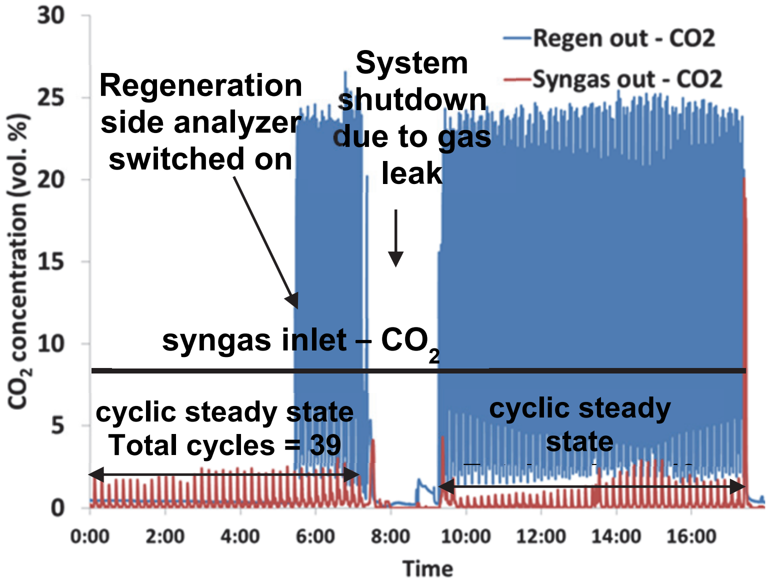


Figure 81. CO_2 Concentrations in the synthesis gas outlet and regeneration streams. CO_2 Capture ~98.7%, Bed Temperature= 200-225°C, P_{ads} =188 psig, P_{des}= 58 psig.

Once the test unit was operating under steady state conditions, different parameters were changed in order to determine what effect they had on the performance of the sorbent.

The parameter that had one of the biggest impacts regarding sorbent capacity was the bed temperature. The average CO_2 capacity (weight of CO_2 adsorbed/weight of sorbent) at the beginning of the field test was slightly lower at 1.2% compared to the baseline testing of 1.6%.

On December 5, 2011, we dropped the bed temperatures by approximately 20°C. This increased the capacities to near 1.6% wt. which matches that of the baseline testing (Figure 82). Figure 83 shows that by decreasing the bed temperature by 20°C increased the capacity by almost 50% without adversely affecting the CO_2 capture percentage.

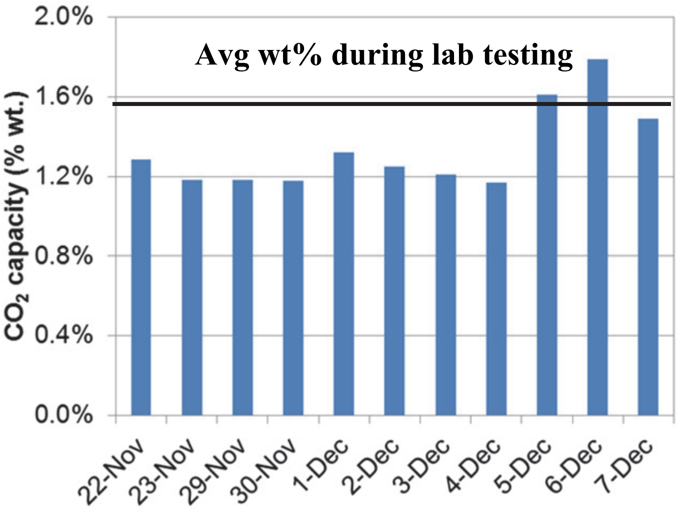


Figure 82 Average CO_2 capacity for duration of testing at the NCCC.

Another operating parameter that was altered was the adsorption time for each bed. While holding all other parameters constant, the adsorption time was varied from 2.7 to 3.05 minutes (Figure 84). As the adsorption time increases, the capacity also increases but the CO_2 removal efficiency decreases (as the sorbent saturates with CO_2, more CO_2 slippage occur). The CO_2 in the packed bed is adsorbed as a wave that travels through the length of the bed. The increase in adsorption time allows for this wave to travel farther through the bed enabling more of the sorbent to be utilized. However, this if this wave is allowed to travel too far, the CO_2 leakage will increase. This is responsible for the decrease in overall CO_2 capture as the adsorption time increases. Increasing the adsorption time for each bed will decrease the overall size of the sorbent beds while increasing the overall efficiency of the entire system.

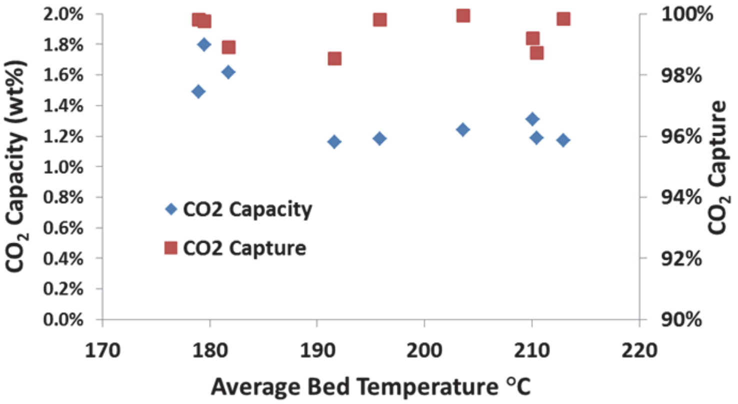


Figure 83. CO_2 capacity related to bed temperature. Syngas flow = 0.4 SCFM, P_{ads}= 178 psig, P_{des}= 44 psig.

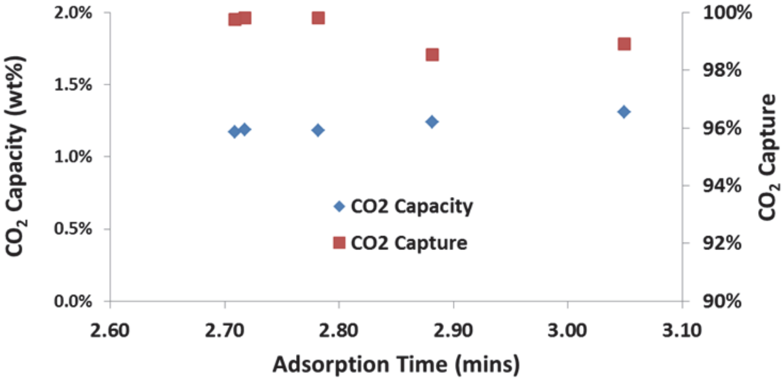


Figure 84. CO_2 capacity and removal efficiency related to adsorption time. Syngas flow 0.4 SCFM P_{ads}= 178 psig, P_{des}= 44 psig.

Longer adsorption times also increases the syngas recovery percentage of system by reducing the number of adsorption and regeneration cycles that a bed must undergo. Therefore, the amount of syngas that is lost during regeneration is reduced.

Once the testing was completed, the Test Unit was brought back to TDA facilities and the initial baseline tests were ran to determine if the performance of the sorbent had changed after being exposed to the syngas contaminates at NCCC and after more than 1000 adsorption cycles. Figure 85 shows that there is no significant degradation in sorbent capacity.

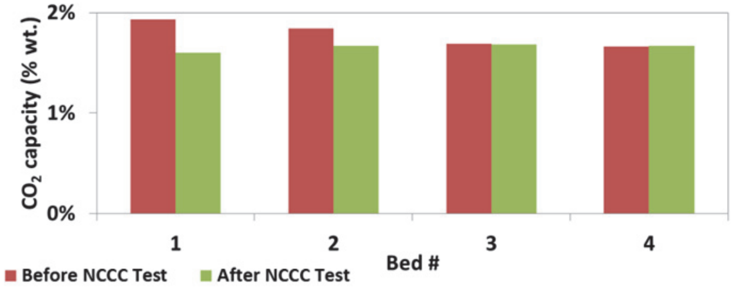


Figure 85. Single bed capacity before and after NCCC testing.

4.10.3 Wabash River Coal Gasification Plant

In August of 2012, the Test Unit was shipped to the Wabash River Coal Gasification Plant in Terre Haute, Indiana (Figure 86). Testing was completed on September 28 2012 at the planned shutdown date of Wabash River IGCC plant. During the testing period, the sorbent underwent over 500 adsorption/regeneration cycles. The sorbent used for this testing is same sorbent that was used during the testing at the National Carbon Capture Center.

Figure 86. Test Unit installed at the Wabash River IGCC Power Plant.

Table 32 provides the gas temperature, pressure and composition of the raw gas. Once the CO is shifted to CO_2 and H_2, the partial pressure of CO_2 was approximately 148 psig.

4.10.4 Wabash River Testing Results

Figure 87 gives a summary of the average performance of the sorbent on each day of testing. The average CO_2 capacity for the testing period is 4.0% wt. and a CO_2 capture percentage of 97.2%. The overall carbon capture of the Test Unit was 95%. This percentage takes into account the carbon that is contained in the unshifted CO that didn't get removed from the synthesis gas stream. This recovery is highly dependent upon the efficacy of the water gas shift reactor ability to convert CO and steam to make H_2 and CO_2.

Table 32 Inlet Gas Stream Compositions at Wabash IGCC Plant

Pressure psig	357	
Temperature C	193	
	Wabash Syngas	Inlet to Sorbent Beds
H_2O	19.9%	5.7%
H_2	23.9%	38.0%
H_2S	1.3%	0.0%
NH_3	0.2%	0.2%
CO	36.1%	10.1%
CH_4	1.6%	1.5%
CO_2	13.8%	41.4%
N_2	2.4%	2.3%
Ar	0.9%	0.8%

The average CO conversion in the water gas shift reactor for the test was 72.9% which is lower than what was seen during the lab testing. This low percentage can be attributed to the rapid flow of syngas in the WGS reactor when a sorbent bed pressurizes during the adsorption step. This rapid flow decreases the residence time in the reactor as well as decreases the steam to CO ratio that is required for the water gas shift reaction thus less CO is shifted. In subsequent tests, a flow restriction and metering device will be in place to regulate the sudden inlet flow into the water gas shift reactor.

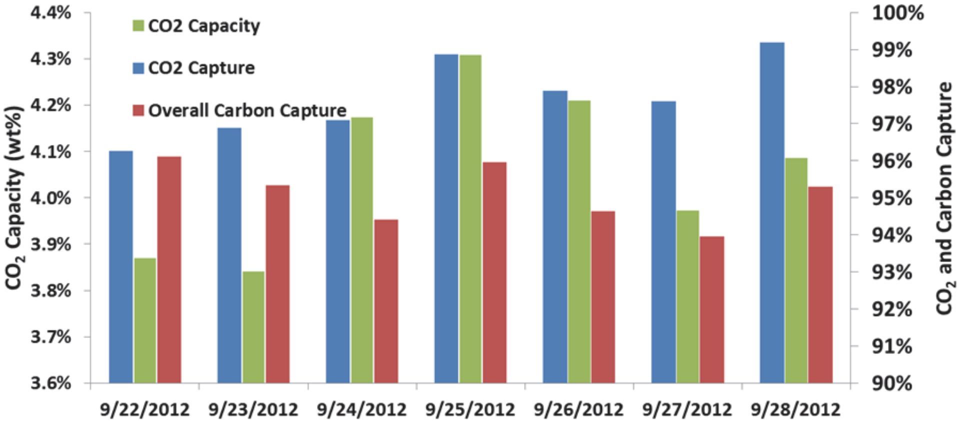


Figure 87. Average sorbent performance for each day of testing.

Figure 88 is a synopsis of the CO_2 concentration in the synthesis gas exiting the test unit while the unit was cycling. There were two system shutdowns, the first was to fix a leaking sample analyzer and the second was to switch to the other sulfur removal bed since the first bed was already spent. One parameter of operation that had the biggest impact on CO_2 leakage was the syngas flow rate. On September 24, the syngas flow out was changed from 0.3 SCFM to 0.5 SCFM. After this increase, the CO_2 leakage increased from 0.5% to a peak of 9%. Later on September 24, the flow was reduced to approximately 0.4 SCFM. Once steady state was achieved, the cycle times were modified in order to reduce the overall CO_2 leakage and boost capacity. The adsorption and regeneration time of the beds was slowly increased from 1 minute to 2 minutes. This increased the CO_2 capture to over 99% and also increased the capacity of sorbent from 4.1% to 4.3% wt.

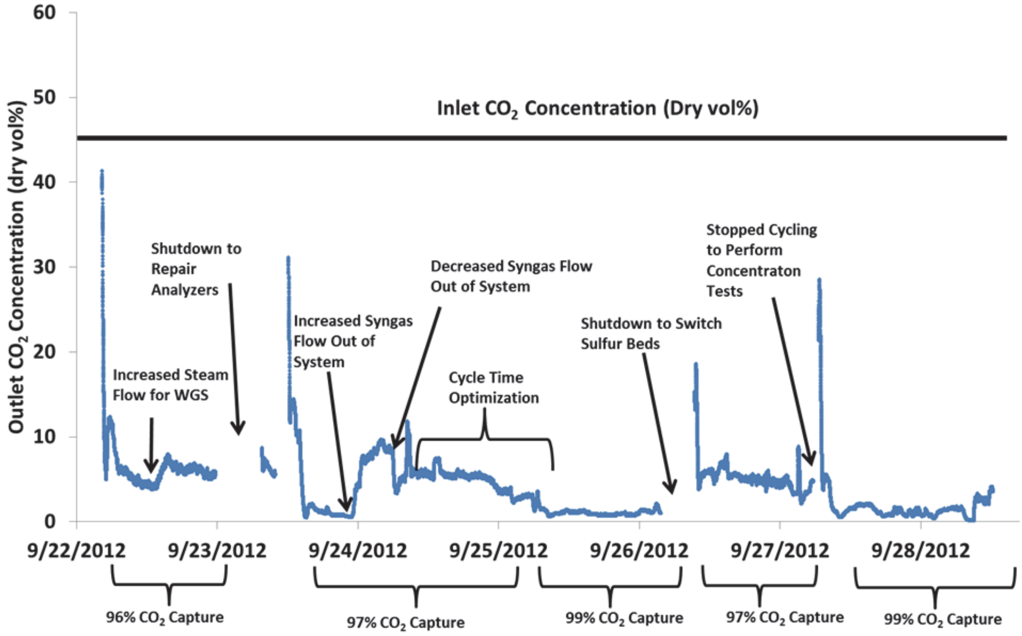


Figure 88. CO_2 outlet concentration for the duration of the Test Period.

The test unit operated with a capture rate of 99% for 24 hours until the unit was stopped for the sulfur bed swap on September 26. Once the unit was back online, the leakage had rose to over 4% without any alterations to the operating parameters. It was discovered later that the shutoff valve on the synthesis gas inlet line had been opened more than during the previous testing. This caused more syngas to enter the water gas shift reactor during sorbent bed pressurization. This dropped the CO conversion due to low residence times in the reactor which resulted in lower a CO_2 partial pressure. This decreases the sorbent capacity and increased the leakage. Flow to the inlet line was reduced later on September 27 and on the 28th, the leakage and capacity were close to previous levels.

The average overall H_2 recovery of the test unit was determined to be 75% during the duration of testing (Figure 89). This is very similar to the recovery numbers that was seen in the

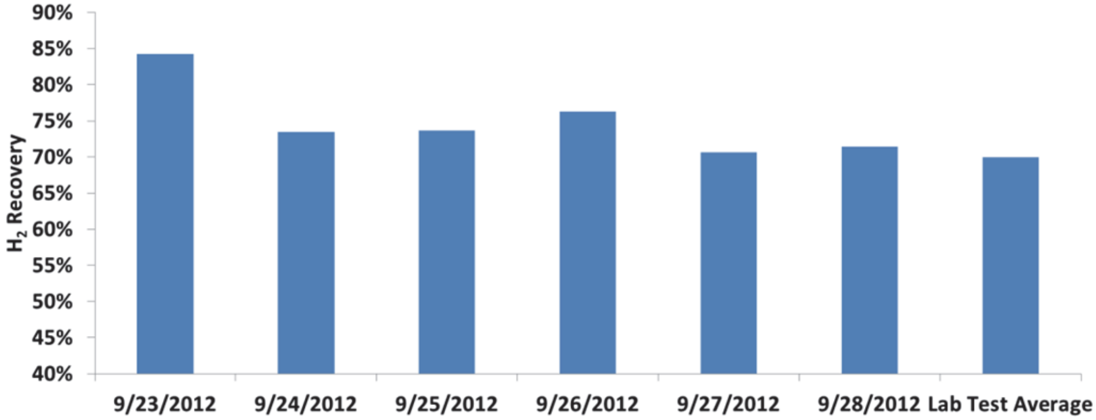


Figure 89. H_2 Recovery during Testing Period in comparison to the laboratory scale tests.

laboratory tests that were performed before the unit was shipped to Wabash River test site. The H_2 recovery is determined by quantifying the amount of H_2 that is lost during regeneration and dividing it by the amount of H_2 that is in the inlet stream. The H_2 recovery during the testing period was 5% more than that seen during the lab testing done at TDA. The H_2 recovery percentages are at slightly lower levels due to the high amounts of dead space volume that is present in small test units such as this one. For example, the ceramic beads located on the top and bottom of each sorbent bed, contribute the dead space of the system. These areas fill up with H_2 during adsorption so during blow down and regeneration, it is readily lost in the regeneration stream. The large scale version of the adsorption unit would forego these high areas of dead space which would help increase its H_2 recovery.

4.10.5 Summary of the Field Testing

Overall, the Test Unit performed as expected when operated with the actual synthesis gas supplied at two separate testing sites (generated by different types of gasifiers running on sub-bituminous coal and petcoke). No degradation of sorbent performance was evident due to the interaction with the synthesis gas contaminants, as confirmed with the test results prior, during and after the field tests. In both demonstrations, the sorbent was tested for over 1,500 adsorption and regeneration cycles using actual synthesis gas without any decrease in its performance.

Figure 90 shows a summary of all the testing. While the Test Unit was at TDA, tests were carried out with the Test Unit simulating the gas composition for the NCCC and Wabash River IGCC Plant. The good match of sorbent performance before and after the field tests suggests that synthesis gas contaminants do not interfere with the operation of the sorbent. As shown in these tests, sorbent capacity is directly related to the CO_2 partial pressure. Based on these results, we project that the sorbent CO_2 capacity will exceed 10% wt. for Phillips 66's next generation gasifier design that will be capable of achieving a CO_2 partial pressure of 240 psi. High CO_2 capacities, resistance to syngas contaminates and its robustness makes TDA's CO_2 sorbent a very viable and economical option for CO_2 sequestration.

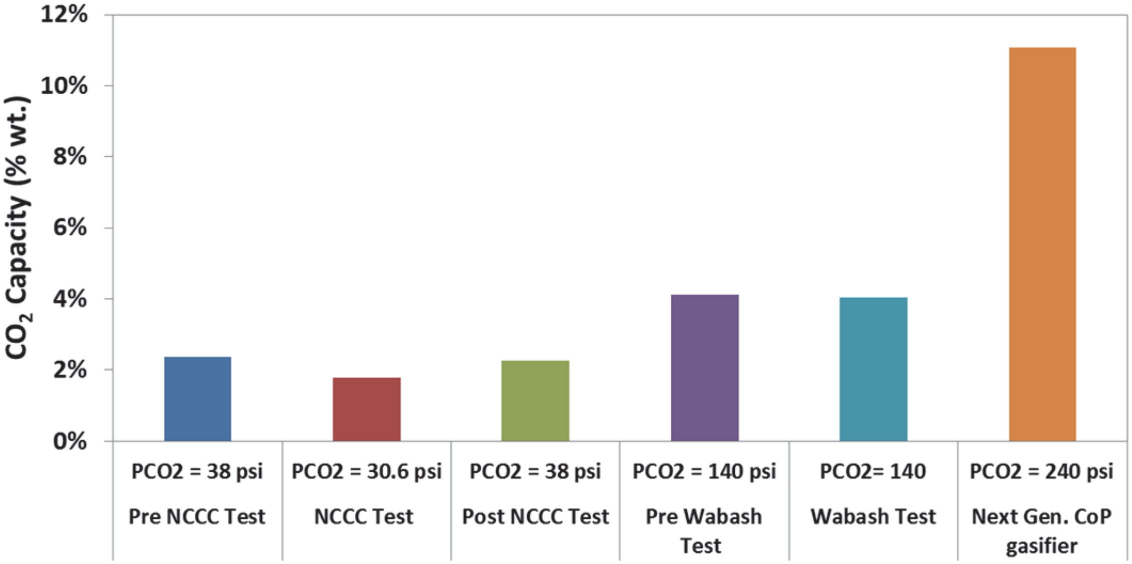


Figure 90. CO_2 capacity of the sorbent for different gasifier demonstrations and projections.

5. Conclusions and Recommendations

In this project we developed a new pre-combustion carbon capture technology and demonstrated its technical feasibility and economic viability in laboratory-scale tests and field demonstrations and carried out a detailed process design and analysis of the new system as part of an Integrated Gasification Combined Cycle (IGCC) power plant. The new technology uses a low cost, high capacity adsorbent that selectively removes CO_2 above the dew point of the synthesis gas (190 to 260°C is explored in this study depending on the gasifier type). The sorbent is based on a TDA proprietary mesoporous carbon that consists of surface functionalized groups that remove CO_2 via physical adsorption. The high surface area and favorable porosity of the sorbent also provide a unique platform to introduce additional functionality, such as active groups to catalyze the water-gas-shift (WGS) reaction or to remove trace metals (e.g., Hg, As).

In collaboration with the Advanced Power and Energy Program of the University of California, Irvine (UCI), TDA developed system simulation models using Aspen Plus^{TM} simulation software to assess the economic viability of TDA's high temperature PSA-based pre-combustion carbon capture technology. The simulation results showed that the CO_2 product purity from TDA's high temperature PSA system at 90% carbon capture could match the purity levels that can be achieved by the Selexol^{TM} process (with an option to further reducing the concentration of Ar and N_2 to less than 20 ppmv, if desired). The TDA's pre-combustion CO_2 capture system achieves a net plant efficiency of 34% on a coal high heating value (HHV) basis. This net plant efficiency is significantly higher than that can be achieved for the Selexol^{TM} scrubber at 31.6%, corresponding to a 7.5% decrease in the heat rate for TDA's Warm Gas Cleanup system. The water consumption in the plant on a net kW generated basis is also significantly lower for the TDA's Warm gas Cleanup process, a reduction of 8.5 kmol per MWh preserving a valuable resource. The plant cost for TDA process is estimated as $2,418/kW, which is 12% lower than that of the Cold Gas Cleanup Case at $2,754/kW. The levelized cost of electricity including the transport, storage and monitoring (TS&M) costs for CO_2 is calculated as $92.9/MWh for TDA's Warm gas Cleanup Case. This cost for the Selexol^{TM} scrubbing technology is calculated as $105.2/MWh.

The results of the system analysis suggest that TDA's high temperature PSA-based Warm Gas Clean-up Technology can make a substantial improvement in the IGCC plant thermal performance for units that are designed to produce near zero emissions, including greater than 90% carbon capture. The capital expenses are also expected to be lower than that of Selexol's™. The higher net plant efficiency and lower capital and operating costs results in substantial reduction in the cost of electricity for the IGCC plant equipped with TDA's high temperature PSA-based carbon capture system.

5.1 Recommendations for Future Work

The results of the DE-FE0000469 project suggest that the high temperature PSA-based pre-combustion carbon capture technology has merits for further research and development. It is recommended that the performance of a fully-equipped system should be demonstrated at larger-scale (e.g., 1 MW) using actual synthesis gas for a longer duration (6 to 12 months). The system should contain all critical components (including all PSA reactors, accumulators etc.) to fully demonstrate the cycle sequence, enabling the demonstration of product purity (both the hydrogen-rich fuel gas and CO_2-rich retentate). The demonstration duration should be long enough to allow over 20,000 cycles (with a 16 min full cycle time it corresponds to 7.5 months of testing) under optimum operating conditions. The selected demonstration scale should be large enough require sorbent production in large quantities (at the recommended 1 MW demonstration 8 m^3 of sorbent will be needed). A more detailed system simulation and cost analysis is also recommended, including design work and accurate quotes from the suppliers of the major process equipment (e.g., air separation unit, gasifier, CO_2 compressors). Successful completion of this recommended work will provide the basis for the new technology to be employed in potential commercial pilot-scale demonstrations (50-100 MW scale).

6. References

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Nomenclature

a_k	film resistance of component k, m/s
B	langmuir parameter, atm^{-1}
D_{ax}	axial dispersion coefficient, m/s^2
D_e/r^2	diffusional time constant, s^{-1}
E	Activation energy, kcal/mol
$-\Delta H$	Heat of adsorption, kcal/mol
$-\Delta H_{ads}$	Heat of adsorption, kcal/mol
$-\Delta H_{des}$	Heat of adsorption, kcal/mol
k_1-k_6	Langmuir-Freundlich temperature dependence parameters
k_b	Langmuir model rate parameter, s^{-1}
k_{b0}	Langmuir model rate constant, s^{-1}
K_0	Henry's constant
L	length of the adsorption column, m
m_t	mass adsorbed at time t, mmol/g
m_∞	mass adsorbed at equilibrium, mmol/g
n	Langmuir-Freundlich constant
P	pressure, atm or bar
P_{CBD}	pressure at the end of countercurrent blowdown step, bar
P_{CO2}	partial pressure of CO_2, atm or bar
P_{DES}	pressure at the end of desorption step, bar
P_H	pressure in the adsorption step, bar
q	adsorbed phase concentration, mmol/g or % wt. CO_2
q_s	adsorbed phase concentration at saturation, mmol/g or % wt. CO_2
q^*	equilibrium adsorbed phase concentration, mmol/g
$\bar{q}$	average adsorbed phase cocentration, mmol/g
q_m	maximum adsorbed phase concentration, mmol/g
R	radius of the spherical particle, m
R_g	Gas constant, kcal/mol/K
t	time, s
T	temperature, K
u	gas phase velocity, m/s
z	axial position in the bed, m
ε	bed porosity due to packing
ε_t	total bed porosity including macropores

Appendices

This section documents the various cases developed prior to the development of the final cases presented in the preceding sections of this report.

Appendix A	Simulation Results to Enhance Carbon Capture for Cold Gas Cleanup Case
Appendix B	Optimization of CO_2 Purification Strategies for Warm Gas Cleanup with TDA's High temperature PSA-based CO_2 capture system
Appendix C	Analysis of TDA'S Warm Gas Cleanup System Using Cryogenic Distillation for CO_2 Purification – Optimization of Adsorption temperature for Phillips 66 and GE gasifiers
Appendix D	Conoco Phillips Gasifier Equipped with Selexol^{TM} Based Cold Gas Cleanup System – Case 4 from DOE/NETL-2010/1397

Appendix A
Simulation Results to Enhance Carbon Capture

A.1 Introduction

A Calibration Case consisting of the Phillips 66 gasifier (Conoco Phillips gasifier when we started the project) and cold gas cleanup was first modeled in Aspen Plus® and the overall thermal performance of the plant was compared to that of a previous DOE/NETL study Case 4 (documented in Report in DOE/NETL-2007/1281 titled, "Cost and Performance Baseline for Fossil Energy Plants," dated August 2007) in order to validate the Aspen Plus® model developed for the entire IGCC system. This model was then modified to reduce the CH_4 content in the gasifier effluent in order to increase the carbon capture from 88.2% for the Calibration Case to 90%. This model was further modified to include the Warm Gas Cleanup while further reducing the CH_4 content in the gasifier effluent to maintain the 90% carbon capture. The performance of the Warm Gas Cleanup Cases were then developed on a consistent basis. In this analysis CO_2 product stream is produced to meet the pipeline specifications listed in Table 20 while limiting the N_2 to 300 ppmv and Ar to 10 ppmv. We used the preliminary PSA system design provided by TDA that was not optimized for the various operating parameters such as operating temperature, regeneration pressure, steam purge volume and the PSA cycle scheme.

A.2 Process Description – Cold Gas Cleanup

The IGCC plant employing the cold gas cleanup and CO_2 capture technology consists of the following plant subsystems:

- Air Separation Unit (ASU)
- Coal Feed Preparation
- Gasification (based on **Phillips 66 E-Gas™ Gasification Technology**)
- High Temperature Syngas Cooling and Scrubbing
- Sour Shifting and Cold Gas Heat Recovery
- Syngas Desulfurization and Decarbonization using a two-stage Selexol™ process
- Claus Sulfur Recovery and Tail Gas Hydrogenation followed by Recycle to the Gasifier
- CO_2 Dehydration and Pressurization (the pressurization scheme was modified to include first compression to a pressure such that the CO_2 stream forms a liquid when cooled against the cooling water, followed by pumping to the final pressure in order to reduce the parasitic load of pressurization)
- Gas Turbine (based on GE F class technology)
- Heat Recovery Steam Generator Unit
- Reheat Steam Cycle

The plant also has the necessary utilities (e.g. cooling water supply, make-up water treatment, plant and instrument air) to support the process units.

Detailed process description of this case may be found in the above referenced DOE/NETL report.

A.3 Process Description - Warm Gas Cleanup

The IGCC plant employing the sorbent CO_2 capture consists of the following plant subsystems:

- Air Separation Unit (ASU)
- Coal Feed Preparation
- Gasification (based on **Phillips 66 E-Gas™ Gasification Technology**)
- High Temperature Syngas Cooling and Scrubbing
- Warm Gas Cleanup similar to RTI's process including Acid (H_2SO_4) Unit
- Sour Shifting
- Regenerable Sorbent CO_2 Capture (based on TDA technology)
- CO_2 Purification (using **Cryogenic Distillation**) and Pressurization
- Gas Turbine (based on GE F class technology)
- Heat Recovery Steam Generator Unit
- Reheat Steam Cycle

The overall configuration of the plant is shown in Figure A-1. Plant subsystems that are different from the Cold Gas Cleanup cases are described in the following.

About 28% of the raw syngas exiting the scrubber is cooled in a heat exchanger while recovering the heat for vacuum condensate / makeup BFW heating followed by trim cooling against cooling water before it is compressed and recycled back to the gasifier system as quench gas.

Remainder of the scrubbed gas is preheated to a temperature of 260°C in a feed/effluent exchanger and supplied to a warm gas cleanup unit similar to RTI's process for removal of sulfur compounds utilizing a zinc titanate adsorbent in a fluidized bed. The performance of this unit as well as the production of H_2SO_4 from the SO_2 in the regenerator off-gas was developed utilizing information available in the public domain. The regenerator off-gas after particulate removal is depressurized by expansion in a power recovery turbine before feeding it to the H_2SO_4 unit. The on-site ASU provides the small amount of O_2 as required by the H_2SO_4 unit in addition to the gasifier. The hot syngas leaving the desulfurizer is cooled to a temperature of 233°C in the feed/effluent exchanger where the desulfurizer feed gas is preheated.

The clean gas is then treated in a sour shift unit similar to the Cold Gas Cleanup cases consisting of two adiabatic beds in series with intercooling where intermediate pressure (IP) and medium pressure (MP) steam is generated. Steam required by the shift unit is supplied as attemperated steam extracted from the steam cycle.

The shifted syngas leaving the last shift reactor at a temperature of 234°C is combined with recycle gas exiting the CO_2 purification unit and then fed to the TDA fixed bed adsorption unit for decarbonizing the syngas before it is combusted in the gas turbines. More than 90% of the syngas enters this decarbonizing unit where 95% of the CO_2 entering with the syngas is separated on a per-pass basis. Remainder of the syngas is utilized in the bed re-pressurization cycle, the adsorbent bed being regenerated at a pressure of 10.34 barA (150 psia). Regeneration is accomplished utilizing 1.31 moles steam per mole CO_2 (roughly 3 reactor volumes of steam purge at 10.34 barA). The mixture consisting of CO_2, steam and residual syngas at a temperature of 208°C is then cooled in a series of heat exchangers while generating low pressure (LP) steam, vacuum condensate / makeup BFW heating and finally trim cooling against cooling water before it is compressed, cooled while recovering bulk of the heat for vacuum condensate / makeup BFW heating, dehydrated and fed to the cryogenic CO_2 purification unit (see Figure A-2). The distillation column operates at a nominal pressure of 52 barA and its primary function is to produce a CO_2 product that meets the pipeline specifications listed in Table 3-2 but limiting the N_2 to 300 ppmv and Ar to 10 ppmv, while minimizing CH_4 losses (dissolved in the liquid CO_2 bottoms stream leaving the column). The bottoms stream is pressurized to the pressure specified in the Table 3-2. The feed to the column is cooled in a series of heat exchangers to a final temperature of -37°C against cold process streams as well as refrigerated liquid propane.

The decarbonized syngas leaving the adsorption unit at a temperature of 265°C with its accompanying unreacted steam is supplied to the gas turbines along with pressurized N_2 from the ASU. The amount of N_2 added to the gas turbine is significantly lower than that in the Cold Gas Cleanup cases, however, due to the large amount of water vapor present in the syngas. The combined cycle design is similar to the design in the Cold Gas Cleanup cases consisting of a reheat steam cycle.

A.4 Cold Gas Cleanup Vs Warm Gas Cleanup

The plant performance of the Calibration Case (which was first modeled in Aspen Plus® to compare its overall plant thermal performance with the DOE/NETL study Case 4 in order to validate the Aspen Plus® IGCC system model) is summarized in Table A-1. The calculated net thermal efficiency of this case is slightly higher at 32.13% versus 31.7% (both on a coal HHV basis) for the DOE/NETL study Case 4. The primary reason for the higher efficiency is that the CO_2 pressurization was accomplished by a combination of gas compression and liquid CO_2 pumping which is more efficient than just gas compression as used in the previous DOE/NETL study. The carbon capture from the syngas for this Calibration Case at 88.2% is limited by the CH_4 content of the syngas.

Phillips 66 can design their gasifier system to produce less CH_4 but at the expense of lower gasifier cold gas efficiency. An additional case with the cold gas cleanup technology was developed with lower CH_4 content in the syngas such that 90% carbon capture may be achieved. The CH_4 content required in the syngas to reach this higher level of carbon capture is 2.86 mole % versus 4.3 mole % (all on a dry basis) for the previous Calibration Case. The performance for this lower CH_4 content case is also summarized in Table A-1. As expected, the overall thermal efficiency dropped, from 32.13% to 31.6% which corresponds to an increase of 1.7% in the heat rate.

The performance of the Warm Gas Cleanup Case is summarized in Table A-2 (which was developed by further modifying the Aspen Plus® IGCC system model developed for the above Cold Gas cleanup Cases). In order to achieve the 90% carbon capture in this Warm Gas Cleanup Case, the CH_4 content in the gasifier effluent had to be further reduced to 2.26 mole % on a dry basis. The resulting efficiency for this case is 33.08% versus 31.6% (both on a coal HHV basis) for the Cold Gas Case when comparing on a consistent carbon capture basis, or an increase of as much as 4.5% in the heat rate. The plant water consumption goes up, however, by almost 9% on a net kW generated basis for the Warm gas Cleanup Case since a significantly higher amount of water vapor enters the gas turbine along with the syngas and is lost to the atmosphere.

The above results indicate that warm gas cleanup technology coupled with the CO_2 adsorption process being developed by TDA can make a substantial improvement in the IGCC plant thermal performance designed for approaching near zero emissions, i.e., to include CO_2 capture.

It should be noted that the coal rate was held constant for each of the three cases while the output of the gas turbine was allowed to vary. Furthermore, the performance of the gas turbine for each of these cases was estimated by UCI and it is recommended that in a more detailed phase of this development program, gas turbine vendors be contacted for providing performance data. The effective LHV of the syngas provided to the gas turbines was held constant for each of the cases by adjusting the amount of diluent, i.e., moisture introduced into the syngas by the humidification operation in the case of cold gas cleanup and amount of N_2 (supplied by the ASU) in the case of the warm gas cleanup. The molar ratio of moisture to N_2 is significantly higher for the warm gas cleanup case and since H_2O has a much higher specific heat than N_2, it may be expected that the NO_x emission for this case will be significantly lower.

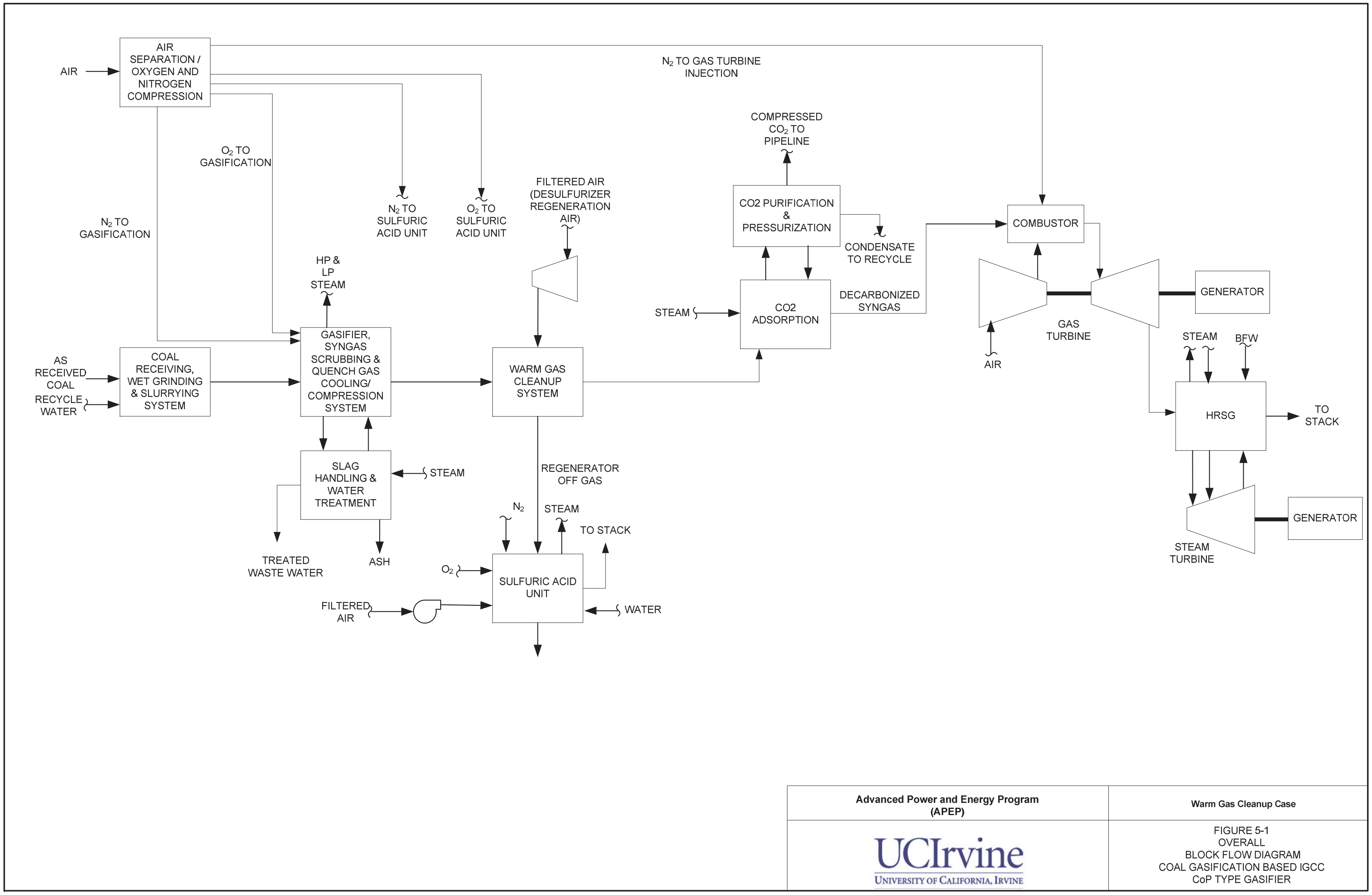


Figure A-1: Block flow diagram for IGCC plant with TDA's high-temperature PSA-based CO_2 capture system

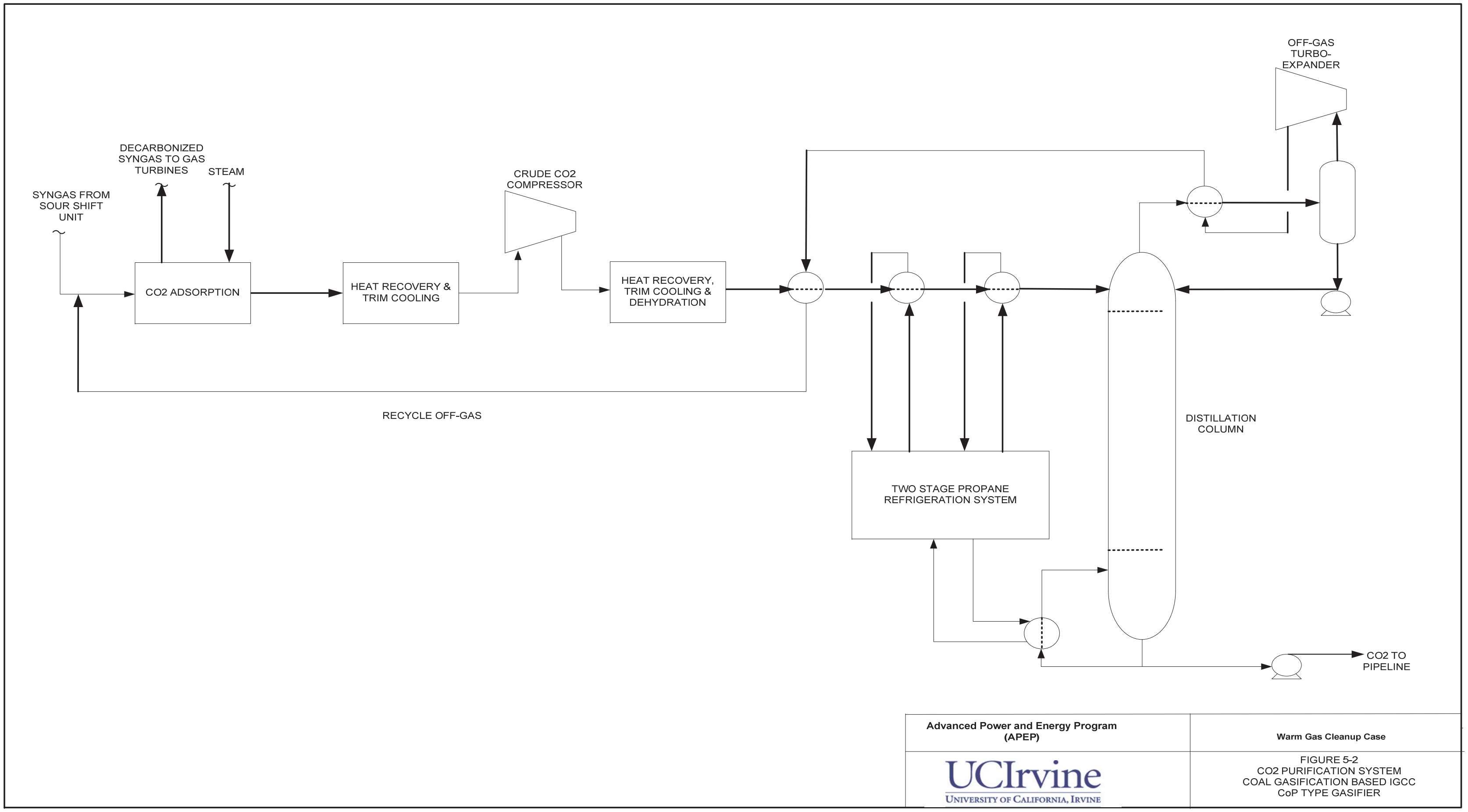
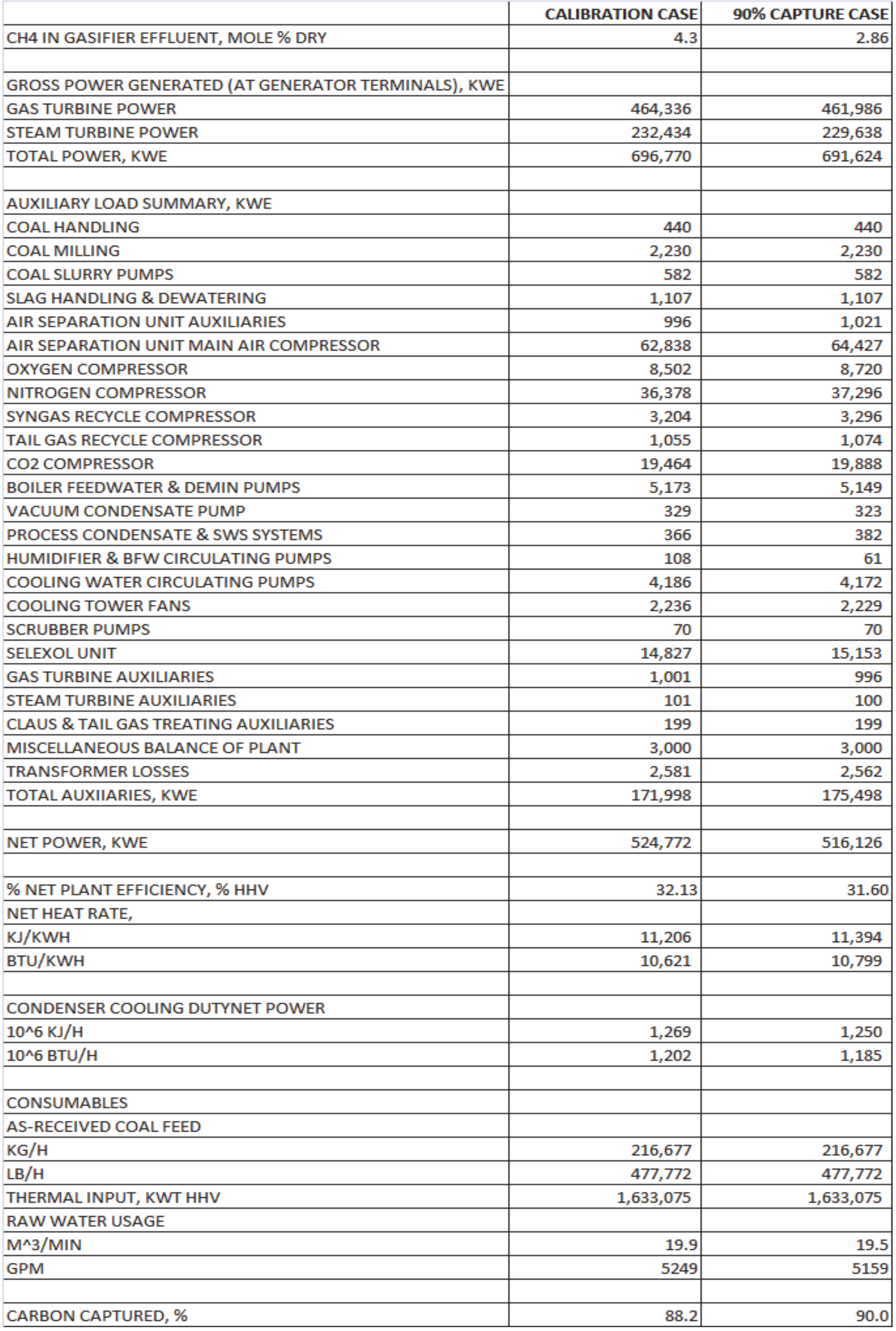


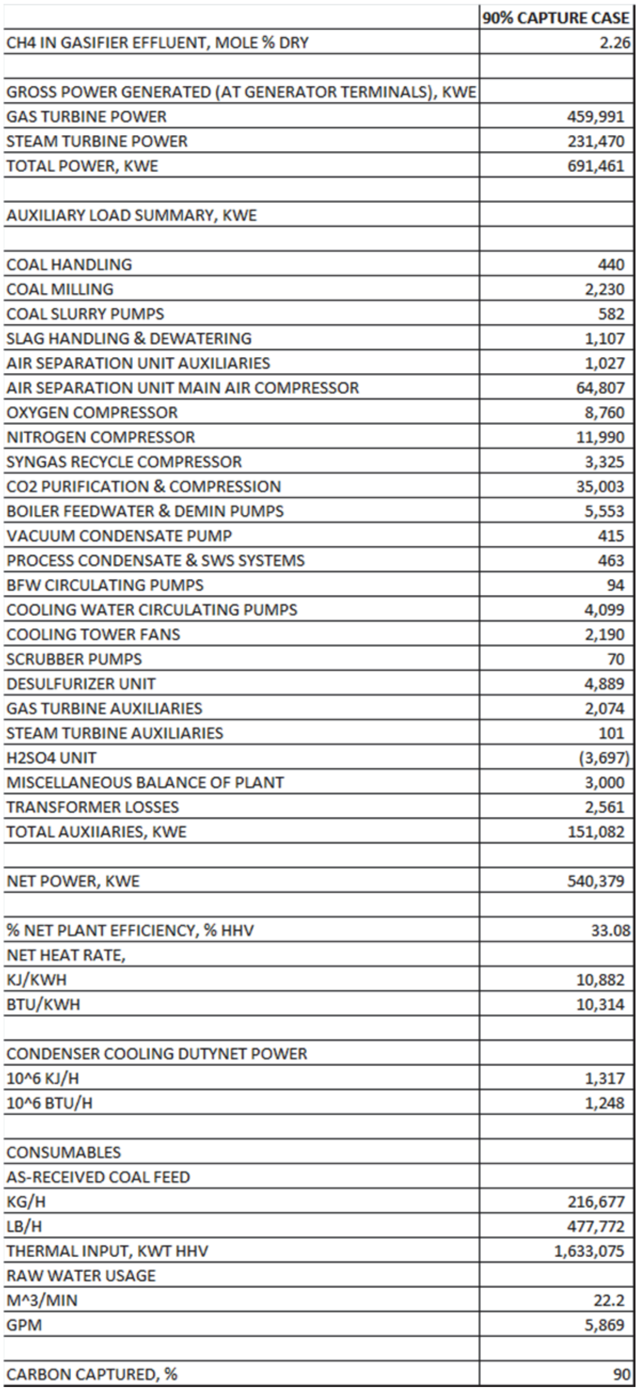
Figure A-2: Block flow diagram of TDA's high temperature PSA-based CO_2 capture system along with the cryogenic CO_2 purification and compression system

Table A-1
Overall Plant Performance Summary
Phase 1 Cold Gas Cleanup Cases
Phillips 66 Gasifier – Constant Coal Feed Rate



	CALIBRATION CASE	90% CAPTURE CASE
CH4 IN GASIFIER EFFLUENT, MOLE % DRY	4.3	2.86
GROSS POWER GENERATED (AT GENERATOR TERMINALS), KWE		
GAS TURBINE POWER	464,336	461,986
STEAM TURBINE POWER	232,434	229,638
TOTAL POWER, KWE	696,770	691,624
AUXILIARY LOAD SUMMARY, KWE		
COAL HANDLING	440	440
COAL MILLING	2,230	2,230
COAL SLURRY PUMPS	582	582
SLAG HANDLING & DEWATERING	1,107	1,107
AIR SEPARATION UNIT AUXILIARIES	996	1,021
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	62,838	64,427
OXYGEN COMPRESSOR	8,502	8,720
NITROGEN COMPRESSOR	36,378	37,296
SYNGAS RECYCLE COMPRESSOR	3,204	3,296
TAIL GAS RECYCLE COMPRESSOR	1,055	1,074
CO2 COMPRESSOR	19,464	19,888
BOILER FEEDWATER & DEMIN PUMPS	5,173	5,149
VACUUM CONDENSATE PUMP	329	323
PROCESS CONDENSATE & SWS SYSTEMS	366	382
HUMIDIFIER & BFW CIRCULATING PUMPS	108	61
COOLING WATER CIRCULATING PUMPS	4,186	4,172
COOLING TOWER FANS	2,236	2,229
SCRUBBER PUMPS	70	70
SELEXOL UNIT	14,827	15,153
GAS TURBINE AUXILIARIES	1,001	996
STEAM TURBINE AUXILIARIES	101	100
CLAUS & TAIL GAS TREATING AUXILIARIES	199	199
MISCELLANEOUS BALANCE OF PLANT	3,000	3,000
TRANSFORMER LOSSES	2,581	2,562
TOTAL AUXIIARIES, KWE	171,998	175,498
NET POWER, KWE	524,772	516,126
% NET PLANT EFFICIENCY, % HHV	32.13	31.60
NET HEAT RATE,		
KJ/KWH	11,206	11,394
BTU/KWH	10,621	10,799
CONDENSER COOLING DUTYNET POWER		
10^6 KJ/H	1,269	1,250
10^6 BTU/H	1,202	1,185
CONSUMABLES		
AS-RECEIVED COAL FEED		
KG/H	216,677	216,677
LB/H	477,772	477,772
THERMAL INPUT, KWT HHV	1,633,075	1,633,075
RAW WATER USAGE		
M^3/MIN	19.9	19.5
GPM	5249	5159
CARBON CAPTURED, %	88.2	90.0

Table A-2
Overall Plant Performance Summary
Phase 1 Warm Gas Cleanup Case
Phillips 66 Gasifier – Constant Coal Feed Rate



	90% CAPTURE CASE
CH4 IN GASIFIER EFFLUENT, MOLE % DRY	2.26
GROSS POWER GENERATED (AT GENERATOR TERMINALS), KWE	
GAS TURBINE POWER	459,991
STEAM TURBINE POWER	231,470
TOTAL POWER, KWE	691,461
AUXILIARY LOAD SUMMARY, KWE	
COAL HANDLING	440
COAL MILLING	2,230
COAL SLURRY PUMPS	582
SLAG HANDLING & DEWATERING	1,107
AIR SEPARATION UNIT AUXILIARIES	1,027
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	64,807
OXYGEN COMPRESSOR	8,760
NITROGEN COMPRESSOR	11,990
SYNGAS RECYCLE COMPRESSOR	3,325
CO2 PURIFICATION & COMPRESSION	35,003
BOILER FEEDWATER & DEMIN PUMPS	5,553
VACUUM CONDENSATE PUMP	415
PROCESS CONDENSATE & SWS SYSTEMS	463
BFW CIRCULATING PUMPS	94
COOLING WATER CIRCULATING PUMPS	4,099
COOLING TOWER FANS	2,190
SCRUBBER PUMPS	70
DESULFURIZER UNIT	4,889
GAS TURBINE AUXILIARIES	2,074
STEAM TURBINE AUXILIARIES	101
H2SO4 UNIT	(3,697)
MISCELLANEOUS BALANCE OF PLANT	3,000
TRANSFORMER LOSSES	2,561
TOTAL AUXIIARIES, KWE	151,082
NET POWER, KWE	540,379
% NET PLANT EFFICIENCY, % HHV	33.08
NET HEAT RATE,	
KJ/KWH	10,882
BTU/KWH	10,314
CONDENSER COOLING DUTYNET POWER	
10^6 KJ/H	1,317
10^6 BTU/H	1,248
CONSUMABLES	
AS-RECEIVED COAL FEED	
KG/H	216,677
LB/H	477,772
THERMAL INPUT, KWT HHV	1,633,075
RAW WATER USAGE	
M^3/MIN	22.2
GPM	5,869
CARBON CAPTURED, %	90

Appendix B
Optimization of CO_2 Purification Strategies for Warm Gas Cleanup with TDA's High temperature PSA-based CO_2 capture system

B.1 Optimization of Purification Schemes for Warm Gas Cleanup

An initial screening analysis was conducted primarily to assess different purification options to produce a CO_2 product stream that meets the pipeline specifications listed in Table 20 while limiting the N_2 to 300 ppmv and Ar to 10 ppmv. The influence of the TDA CO_2 adsorption process regeneration pressure on the overall IGCC plant performance was also assessed however the steam purge volume was also simultaneously reduced to 2.4 bed volumes. The results of this analysis are summarized in Table B-1. The results indicated a cryogenic distillation process is required to meet the N_2 and Ar limits in the CO_2 product and lower adsorbent regeneration pressure below 10.34 barA did not show any advantage.

Table B-1
Overall Plant Performance Summary
Phase 1 Warm Gas Cleanup Screening Cases
Phillips 66 Gasifier

Case #	Regeneration		Recycle			CO_2 Product			Plant Net Thermal Efficiency
	Steam Supply/Use Pressures, psia	Steam Flow Rate, moles H_2O/mole CO_2 adsorbed	Recycle Rate	CO_2 Concentration in Dry Gas to Purification, mole % based on data provided by TDA (sets recycle rate since syngas components leaving with desorbed CO_2 end up being recycled from the purification step)	Purification	Specification	Per Pass CO_2 Capture, %	Overall Carbon Capture, % (using Bypass Control)	% Coal HHV
Case 1-1	High (150/145.1)	14,746/11,221.78 = 1.31	Higher than Phase 1	85	Cryogenic Separation (Distillation)	Per Table 3-2 but with the stringent N_2 and Ar specifications	98.38	90.00	32.60
Case 1-2	High (150/145.1)	1.31	None	85	Cryogenic Separation (Distillation) + Catalytic Combustion of Impurities (Distillate)	Per Table 3-2	98.38	90.00	Tabled as it shows no advantage over Case 1-4. Distillate (containing 68% H_2 by vol) combustor exit temperature excessive (> 3400°C) while using near stoichmetric O_2 (in order to minimize its usage).
Case 1-3	High (150/145.1)	1.31	Minimum (Impurities from Cryogenic separation)	85	Catalytic Combustion of Raw CO_2 followed by Cryogenic Separation (Distillate)	Per Table 3-2 but with the stringent N_2 and Ar specifications	98.38	90.00	32.54
Case 1-4	High (150/145.1)	1.31	None based on update shown for purification	85	Catalytic Combustion only, no Cryogenic Separation	Per Table 3-2	98.38	90.00	32.79
Case 1-5	Low (70/65.3)	7,268/10,858.66 = 0.6693	Minimum (Impurities from Cryogenic separation)	89.3	Catalytic Combustion of Raw CO_2 followed by Cryogenic Separation (Distillate)	Per Table 3-2 but with the stringent N_2 and Ar specifications	95.64	90.00	32.09

Appendix C
Analysis of TDA'S Warm Gas Cleanup System Using Cryogenic Distillation for CO_2 Purification – Optimization of Adsorption temperature for Phillips 66 and GE gasifiers

C.1 Introduction

In this study, the TDA technology was evaluated in both the Phillips 66 gasifier and the General Electric gasifier while utilizing updated performance and costs for the cold gas cleanup cases documented as Case 4 and Case 2, respectfully in the DOE NETL study report DOE/NETL-2010/1397 titled, "Cost and Performance Baseline for Fossil Energy Plants," dated November 2010. The calibration cases were again modeled in Aspen Plus® and the overall thermal performances of the plants were compared to those of the DOE/NETL study Case 4 and Case 2 in order to validate the Aspen Plus® model developed for the two entire IGCC systems. The performance and costs of the Warm Gas Cleanup Cases were then developed on a consistent basis using updated PSA cycle scheme and system design from TDA where the steam purge volume is further optimized to 2.4 reactor volumes. We also explored the impact of adsorption temperature on the net plant efficiency in the warm gas capture cases with both the Phillips 66 and GE gasifiers.

C.2 Process Description - Cold Gas Cleanup Cases

The IGCC plants employing the cold gas cleanup and CO_2 capture technology consists of the following plant subsystems:

- Air Separation Unit (ASU)
- Coal Feed Preparation
- Gasification (based on **Phillips 66 Technology** and **General Electric Technology**)
- High Temperature Syngas Cooling and Scrubbing
- Sour Shifting and Cold Gas Heat Recovery
- Syngas Desulfurization and Decarbonization using a two-stage Selexol™ process
- Claus Sulfur Recovery and Tail Gas Hydrogenation followed by Recycle to the Gasifier
- CO_2 Dehydration and Pressurization (the pressurization scheme was modified to include first compression to a pressure such that the CO_2 stream forms a liquid when cooled against the cooling water, followed by pumping to the final pressure in order to reduce the parasitic load of pressurization)
- Gas Turbine (based on GE F class technology)
- Heat Recovery Steam Generator Unit
- Reheat Steam Cycle

The plant also has the necessary utilities (e.g. cooling water supply, make-up water treatment, plant and instrument air) to support the process units.

Detailed process description of this case may be found in the above referenced DOE/NETL report.

C.3 Process Description - Warm Gas Cleanup Cases

The IGCC plants employing the sorbent CO_2 capture consist of the following plant subsystems:

- Air Separation Unit (ASU)
- Coal Feed Preparation
- Gasification (based on **Phillips 66 Technology** and **GE Technology**)
- High Temperature Syngas Cooling and Scrubbing
- Warm Gas Cleanup similar to RTI’s process including Acid (H_2SO_4) Unit
- Sour Shifting
- Regenerable Sorbent CO_2 Capture (based on TDA technology)
- CO_2 Purification (using **Distillation**) and Pressurization
- Gas Turbine (based on GE F class technology)
- Heat Recovery Steam Generator Unit
- Reheat Steam Cycle

The plants also have the necessary utilities (e.g. cooling water supply, make-up water treatment, plant and instrument air) to support the process units.

Two different feed gas temperatures for the syngas entering the sorbent CO_2 capture process were evaluated in the case of Phillips 66 gasifier, one enters at 198°C and the other enters at 240°C (35°C and 77°C above the dew point of the synthesis gas respectively) while in the case of GE gasifier, one enters at 215°C and the other enters at 258°C (33°C and 76 above the dew point of the synthesis gas respectively).

C.4 Cold Gas Vs Warm Gas Cleanup

C.4.1 Phillips 66 Gasifier

The impact of changing the syngas feed temperature to the CO_2 adsorption (PSA) unit in the Warm Gas Cleanup was studied, in addition to making a comparison with the Cold Gas Cleanup, all on a consistent basis. The net efficiency for the Cold Gas Case as presented in Table C-1 at 31.62% is significantly lower than those for the Warm Gas Cleanup Cases as presented in Table C-2. The corresponding efficiencies for the two syngas inlet temperatures of 198°C and 240°C are 33.49% and 33.28% (on a coal HHV basis), respectfully.

The plant cost and the levelized cost of electricity estimates for the Cold Gas Cleanup Case and the Warm gas Cleanup Cases with the two syngas inlet temperatures of 198°C and 240°C are presented in Tables C-3 through C-6. The resulting plant costs for the Warm gas Cleanup Cases are $2,537/kW and $2,567/ kW while that for the Cold Gas Case is $2,754/kW, again a significant reduction in cost for the Warm Gas Cleanup Cases. The levelized cost of electricity with transport, storage and monitoring (TS&M) costs for the CO_2 included is $96.6/MWh and $98.0/MWh for the two Warm gas Cleanup Cases with syngas inlet temperatures of 198°C and 240°C. This shows that there is a significant advantage in lowering the inlet synthesis feed gas temperature to the CO_2 adsorption (PSA) unit to 35°C above the dew point. The primary reason for the lower COE is due to the decrease in the amount of adsorbent required for lower temperature. The corresponding levelized COE for the Cold Gas Cleanup Case is $105.2/MWh which is significantly higher than those for the Warm Gas Cleanup Cases.

C.4.2 Phillips 66 Gasifiers

The impact of changing the syngas feed temperature to the CO_2 adsorption (PSA) unit in the Warm Gas Cleanup was again studied this time for GE gasifier based IGCC plant, in addition to making a comparison with the Cold Gas Cleanup, all on a consistent basis. The net efficiency for the Cold Gas Case with GE gasifier as presented in Table C-7 at 32.38% is again lower than those for the Warm Gas Cleanup Cases as presented in Table C-8. The corresponding efficiencies for the two syngas inlet temperatures of 215°C and 258°C are 34.19% and 33.85% (on a coal HHV basis), respectfully.

Again, lower syngas temperature for the Warm Gas Cleanup showed significant improvement in efficiency, the economic analysis was performed only on the lower temperature case. The plant cost and the levelized cost of electricity estimates for the Cold Gas Cleanup Case and the Warm gas Cleanup Case are presented in Tables C-9 through C-12. The resulting plant cost for the Warm gas Cleanup Case is $2,628/kW while that for the Cold Gas Case is $2,703/kW, again a reduction in cost for the Warm Gas Cleanup Cases. The levelized cost of electricity with transport, storage and monitoring (TS&M) costs for the CO_2 included is $98.5/MWh for the Warm Gas Cleanup. The corresponding levelized COE for the Cold Gas Cleanup is $103.0/MWh which is again higher than that for the Warm Gas Cleanup.

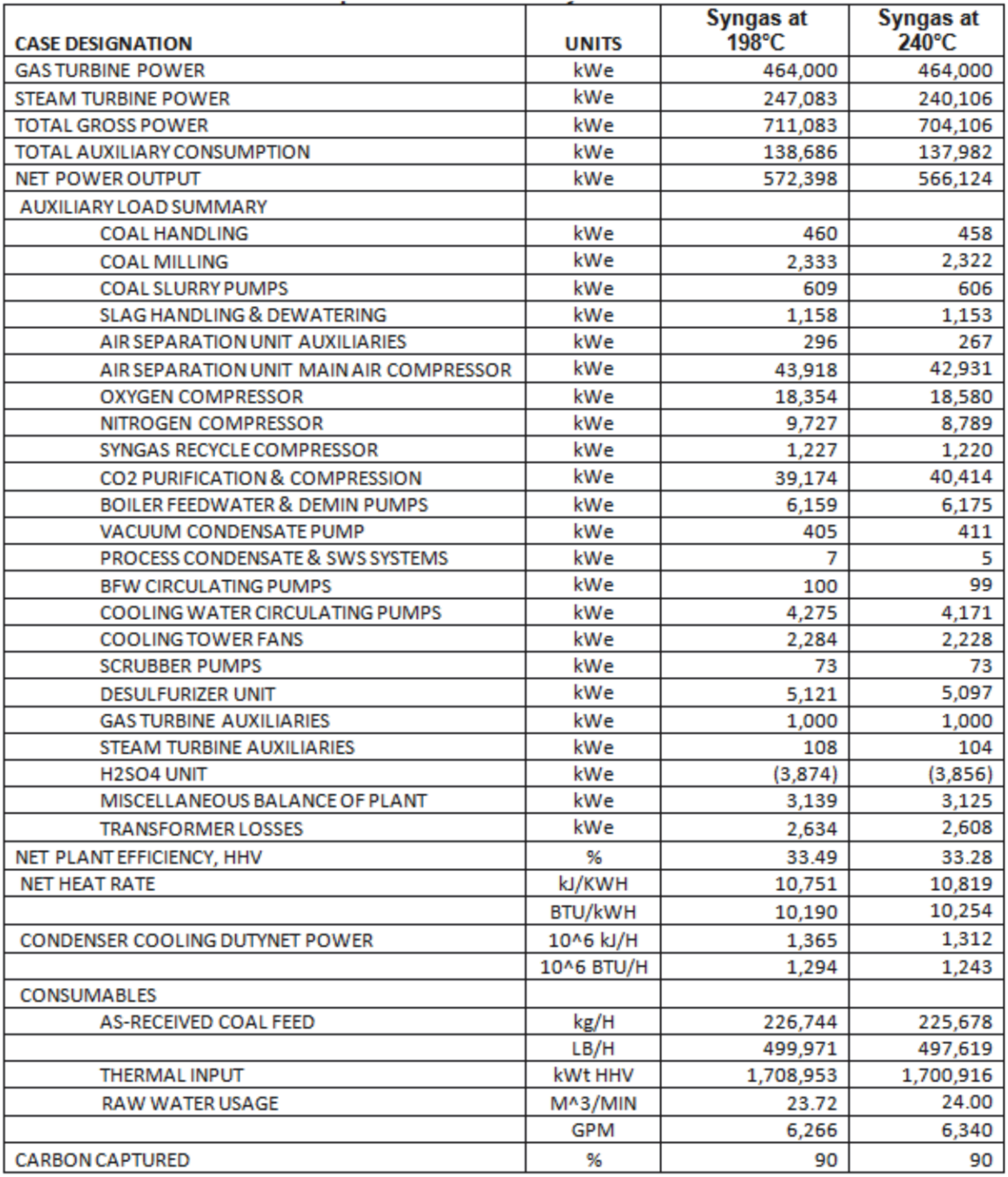
C.4.3 Avoided CO_2 Costs

The cost of capturing the CO_2 on a per tonne of CO_2 avoided is calculated for both the Phillips 66 and the General Electric gasifier cases are presented in Table C-13. As can be seen, again, the Warm Gas Cleanup Cases show the lowest avoided costs of capturing the CO_2.

Table C-1
Overall Plant Performance Summary – Phase 2 (Updated) Cold Gas Cleanup Case
Phillips 66 Gasifier – Fully Loaded GTs

	UNITS	
GAS TURBINE POWER	kWe	464,000
STEAM TURBINE POWER	kWe	227,247
TOTAL POWER	kWe	691,247
TOTAL AUXILIARY CONSUMPTION	kWe	175,994
NET POWER OUTPUT	kWe	515,253
AUXILIARY LOAD SUMMARY	kWe	
COAL HANDLING	kWe	439
COAL MILLING	kWe	2,225
COAL SLURRY PUMPS	kWe	581
SLAG HANDLING & DEWATERING	kWe	1,159
AIR SEPARATION UNIT AUXILIARIES	kWe	1,039
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	kWe	63,739
OXYGEN COMPRESSOR	kWe	8,873
NITROGEN COMPRESSOR	kWe	34,284
SYNGAS RECYCLE COMPRESSOR	kWe	1,117
TAIL GAS RECYCLE COMPRESSOR	kWe	3,190
CO2 COMPRESSOR	kWe	19,849
BOILER FEEDWATER & DEMIN PUMPS	kWe	5,723
VACUUM CONDENSATE PUMP	kWe	353
PROCESS CONDENSATE & SWS SYSTEMS	kWe	299
HUMIDIFIER & BFW CIRCULATING PUMPS	kWe	44
COOLING WATER CIRCULATING PUMPS	kWe	4,286
COOLING TOWER FANS	kWe	2,290
SCRUBBER PUMPS	kWe	70
SELEXOL UNIT	kWe	19,582
GAS TURBINE AUXILIARIES	kWe	1,000
STEAM TURBINE AUXILIARIES	kWe	99
CLAUS & TAIL GAS TREATING AUXILIARIES	kWe	199
MISCELLANEOUS BALANCE OF PLANT	kWe	2,993
TRANSFORMER LOSSES	kWe	2,560
NET PLANT EFFICIENCY, HHV	%	31.62
NET HEAT RATE	kJ/kWH	11,387
	BTU/kWH	10,793
CONDENSER COOLING DUTY	10^6 kJ/H	1,254
	10^6 BTU/H	1,189
CONSUMABLES		
AS-RECEIVED COAL FEED	kg/H	216,187
	LB/H	476,692
THERMAL INPUT	kWt HHV	1,629,385
RAW WATER USAGE	M^3/MIN	23.07
	GPM	6,094
CARBON CAPTURED	%	90

Table C-2
Overall Plant Performance Summary – Phase 2 Warm Gas Cleanup Cases
Phillips 66 Gasifier – Fully Loaded GTs



CASE DESIGNATION	UNITS	Syngas at 198°C	Syngas at 240°C
GAS TURBINE POWER	kWe	464,000	464,000
STEAM TURBINE POWER	kWe	247,083	240,106
TOTAL GROSS POWER	kWe	711,083	704,106
TOTAL AUXILIARY CONSUMPTION	kWe	138,686	137,982
NET POWER OUTPUT	kWe	572,398	566,124
AUXILIARY LOAD SUMMARY			
COAL HANDLING	kWe	460	458
COAL MILLING	kWe	2,333	2,322
COAL SLURRY PUMPS	kWe	609	606
SLAG HANDLING & DEWATERING	kWe	1,158	1,153
AIR SEPARATION UNIT AUXILIARIES	kWe	296	267
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	kWe	43,918	42,931
OXYGEN COMPRESSOR	kWe	18,354	18,580
NITROGEN COMPRESSOR	kWe	9,727	8,789
SYNGAS RECYCLE COMPRESSOR	kWe	1,227	1,220
CO2 PURIFICATION & COMPRESSION	kWe	39,174	40,414
BOILER FEEDWATER & DEMIN PUMPS	kWe	6,159	6,175
VACUUM CONDENSATE PUMP	kWe	405	411
PROCESS CONDENSATE & SWS SYSTEMS	kWe	7	5
BFW CIRCULATING PUMPS	kWe	100	99
COOLING WATER CIRCULATING PUMPS	kWe	4,275	4,171
COOLING TOWER FANS	kWe	2,284	2,228
SCRUBBER PUMPS	kWe	73	73
DESULFURIZER UNIT	kWe	5,121	5,097
GAS TURBINE AUXILIARIES	kWe	1,000	1,000
STEAM TURBINE AUXILIARIES	kWe	108	104
H2SO4 UNIT	kWe	(3,874)	(3,856)
MISCELLANEOUS BALANCE OF PLANT	kWe	3,139	3,125
TRANSFORMER LOSSES	kWe	2,634	2,608
NET PLANT EFFICIENCY, HHV	%	33.49	33.28
NET HEAT RATE	kJ/KWH	10,751	10,819
	BTU/kWH	10,190	10,254
CONDENSER COOLING DUTYNET POWER	10^6 kJ/H	1,365	1,312
	10^6 BTU/H	1,294	1,243
CONSUMABLES			
AS-RECEIVED COAL FEED	kg/H	226,744	225,678
	LB/H	499,971	497,619
THERMAL INPUT	kWt HHV	1,708,953	1,700,916
RAW WATER USAGE	M^3/MIN	23.72	24.00
	GPM	6,266	6,340
CARBON CAPTURED	%	90	90

Table C-3
Overall Plant Cost Summary
Phase 2 Cold Gas Cleanup Case
Phillips 66 Gasifier – Fully Loaded GTs

UNIT	2007 Installed Cost ($1000)
ASU	220,067
Fuel receiving, preparation & feeding	92,151
Gasifier, syngas cooler & aux	248,837
Gasification foundations	18,067
Ash handling systems	37,123
Flare stack system	3,008
Shift reactor	13,534
Syngas scrubber (included in gasification auxiliaries)	0
Blowback gas systems	1,330
Fuel gas piping	1,574
Gas cleanup foundations	1,703
Hg Removal + LTGC	46,712
Selexol	199,729
Claus + TG Recycle	33,288
CO2 compression, dehydration + pumping	24,586
Gas turbine + generator + auxiliaries	132,015
HRSG, ducting + stack	56,119
Steam turbine + generator + auxiliaries	50,719
Surface condenser	6,984
Feedwater system	19,616
Water makeup + pretreating	2,035
Other feedwater subsystems	3,208
Service water systems	5,842
Other boiler plant systems	6,468
Fuel oil system & nat gas	1,915
Waste water treatment	2,251
Misc. power plant equipment	2,537
Cooling water system	35,313
Accessory electric plant	87,698
Instrumentationo & controls	27,099
Improvement to site	19,578
Buildings & structures	17,936
Total	1,419,040
Total $/kW	**2,754**

Table C-4
Overall Plant Cost Summary
Warm Gas Cleanup Cases
Phillips 66 Gasifier – Fully Loaded GTs

UNIT	2007 Installed Cost ($1000)	
Syngas Temperature	**198°C**	**240°C**
ASU	228,079	228,766
Fuel receiving, preparation & feeding	95,278	94,964
Gasifier, syngas cooler & aux	257,282	256,435
Gasification foundations	18,680	18,618
Ash handling systems	38,382	38,256
Flare stack system	3,110	3,099
Warm gas desulfurization	24,826	24,435
H2SO4 unit	54,240	53,387
Shift reactor	13,398	13,348
Blowback gas systems	1,375	1,371
Fuel gas piping	2,314	2,335
Gas cleanup foundations	1,685	1,679
Hg Removal, CO2 separation	111,812	113,969
CO2 purification / refrigeration + pumping	111,543	113,695
Gas turbine + generator + auxiliaries	132,015	132,015
HRSG, ducting + stack	57,973	58,040
Steam turbine + generator + auxiliaries	53,779	52,711
Surface condenser	7,410	7,207
Feedwater system	22,451	23,771
Water makeup + pretreating	2,075	2,092
Other feedwater subsystems	3,401	3,334
Service water systems	6,040	6,020
Other boiler plant systems	6,858	6,722
Fuel oil system & nat gas	1,915	1,915
Waste water treatment	2,327	2,320
Misc. power plant equipment	2,588	2,570
Cooling water system	35,248	34,647
Accessory electric plant	89,452	88,837
Instrumentationo & controls	28,019	27,926
Improvement to site	20,242	20,176
Buildings & structures	18,545	18,484
Total	1,452,342	1,453,145
Total $/kW	**2,537**	**2,567**

Table C-5
Cost of Electricity
Cold Gas Cleanup Case
Phillips 66 Gasifier – Fully Loaded GTs
2007 $

Net power, MW	515.25
Capacity factor (CF), %	80
Total plant cost (TPC), $	1,419,040,277
6 month labor cost	12,883,919
1 month maintenance materials	2,213,213
1 month non-fuel consumables	531,751
1 month waste disposal	290,027
25% of 1 month fuel cost at 100% CF	1,660,760
2% of TPC	28,380,806
60 day supply of fuel & consumables at 100% CF	14,153,013
0.5% of TPC (spare parts)	7,095,201
Initial catalyst & chemicals cost, $	7,180,117
Land	900,000
Other owners's costs (15% of TPC)	212,856,041
Financing costs	38,314,087
Total overnight cost (TOC), $	1,745,499,213
Fixed operating cost for initial year of operation (OCF), $	54,148,643
Annual feed cost at above CF for initial year (OCV1), $	63,773,186
Other annual variable operating cost at above CF for initial year (OCV2), $	29,135,919
Annual CO2 transporting, storing, and monitoring cost at above CF for initial year (OCV3), $	19,419,868
Annual byproduct revenues at above CF for initial year (OCV4), $	3,749,093
1st year cost of electricity (COE) w/o CO2 TS&M, $/MWh	**99.8**
1st year cost of electricity (COE), $/MWh	**105.2**

Table C-6
Cost of Electricity
Warm Gas Cleanup Cases
Phillips 66 Gasifier – Fully Loaded GTs
2007 $

Syngas Temperature	**198°C**	**240°C**
Net power, MW	572.40	566.12
Capacity factor (CF), %	80	80
Total plant cost (TPC), $	1,452,342,140	1,453,144,776
6 month labor cost	13,115,595	13,109,975
1 month maintenance materials	2,270,581	2,269,189
1 month non-fuel consumables	910,738	1,021,714
1 month waste disposal	372,994	390,513
25% of 1 month fuel cost at 100% CF	1,741,860	1,733,668
2% of TPC	29,046,843	29,062,896
60 day supply of fuel & consumables at 100% CF	15,540,517	15,694,791
0.5% of TPC (spare parts)	7,261,711	7,265,724
Initial catalyst & chemicals cost, $	12,416,487	15,018,943
Land	900,000	900,000
Other owners's costs	217,851,321	217,971,716
Financing costs	39,213,238	39,234,909
Total overnight cost (TOC), $	1,792,984,025	1,796,818,814
Fixed operating cost for initial year of operation (OCF), $	55,278,033	55,282,845
Annual feed cost at above CF for initial year (OCV1), $	66,887,435	66,572,864
Other annual variable operating cost at above CF for initial year (OCV2), $	34,121,403	35,341,590
Annual CO2 transporting, storing, and monitoring cost at above CF for initial year (OCV3), $	20,368,203	20,272,411
Annual byproduct revenues at above CF for initial year (OCV4), $	12,094,419	12,047,942
1st year cost of electricity (COE) w/o CO2 TS&M, $/MWh	**91.5**	**92.9**
1st year cost of electricity (COE), $/MWh	**96.6**	**98.0**

Table C-7
Overall Plant Performance Summary
Phase 2 Cold Gas Cleanup Case
General Electric Gasifier – Fully Loaded GTs

	UNITS	
GAS TURBINE POWER	kWe	464,000
STEAM TURBINE POWER	kWe	245,169
SYNGAS EXPANDER POWER		6,502
TOTAL POWER	kWe	715,671
TOTAL AUXILIARY CONSUMPTION	kWe	176,380
NET POWER OUTPUT	kWe	539,291
AUXILIARY LOAD SUMMARY		
COAL HANDLING	kWe	449
COAL MILLING	kWe	2,273
COAL SLURRY PUMPS	kWe	1,048
SLAG HANDLING & DEWATERING	kWe	1,128
AIR SEPARATION UNIT AUXILIARIES	kWe	1,089
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	kWe	67,405
OXYGEN COMPRESSOR	kWe	10,676
NITROGEN COMPRESSOR	kWe	35,694
TAIL GAS RECYCLE COMPRESSOR	kWe	1,212
CO2 COMPRESSOR	kWe	20,016
BOILER FEEDWATER & DEMIN PUMPS	kWe	4,498
VACUUM CONDENSATE PUMP	kWe	321
PROCESS CONDENSATE & SWS SYSTEMS	kWe	780
HUMIDIFIER & BFW CIRCULATING PUMPS	kWe	30
COOLING WATER CIRCULATING PUMPS	kWe	4,815
COOLING TOWER FANS	kWe	2,572
SCRUBBER PUMPS	kWe	71
SELEXOL UNIT	kWe	15,283
GAS TURBINE AUXILIARIES	kWe	1,000
STEAM TURBINE AUXILIARIES	kWe	107
CLAUS & TAIL GAS TREATING AUXILIARIES	kWe	203
MISCELLANEOUS BALANCE OF PLANT	kWe	3,059
TRANSFORMER LOSSES	kWe	2,651
NET PLANT EFFICIENCY, HHV	%	32.38
NET HEAT RATE	kJ/kWH	11,117
	BTU/kWH	10,537
CONDENSER COOLING DUTY	10^6 kJ/H	1,517
	10^6 BTU/H	1,438
CONSUMABLES		
AS-RECEIVED COAL FEED	kg/H	220,906
	LB/H	487,097
THERMAL INPUT	kWt HHV	1,664,950
RAW WATER USAGE	M^3/MIN	21.20
	GPM	5,601
CARBON CAPTURED	%	90

Table C-8
Overall Plant Performance Summary – Phase 2 Warm Gas Cleanup Cases
General Electric Gasifier – Fully Loaded GTs

CASE DESIGNATION	**UNITS**	**Syngas at 215°C**	**Syngas at 258°C**
GAS TURBINE POWER	kWe	464,000	464,000
STEAM TURBINE POWER	kWe	253,622	248,946
SYNGAS EXPANDER POWER	kWe	11,623	11,769
TOTAL GROSS POWER	kWe	729,245	736,485
TOTAL AUXILIARY CONSUMPTION	kWe	135,424	136,298
NET POWER OUTPUT	kWe	593,864	588,418
AUXILIARY LOAD SUMMARY			
COAL HANDLING	kWe	468	468
COAL MILLING	kWe	2,371	2,373
COAL SLURRY PUMPS	kWe	1,093	1,094
SLAG HANDLING & DEWATERING	kWe	1,176	1,177
AIR SEPARATION UNIT AUXILIARIES	kWe	150	120
AIR SEPARATION UNIT MAIN AIR COMPRESSOR	kWe	43,313	42,534
OXYGEN COMPRESSOR	kWe	20,010	20,301
NITROGEN COMPRESSOR	kWe	5,333	4,297
CO2 PURIFICATION & COMPRESSION	kWe	40,010	42,331
BOILER FEEDWATER & DEMIN PUMPS	kWe	5,343	5,355
VACUUM CONDENSATE PUMP	kWe	358	370
PROCESS CONDENSATE & SWS SYSTEMS	kWe	633	689
BFW CIRCULATING PUMPS	kWe	102	102
COOLING WATER CIRCULATING PUMPS	kWe	4,389	4,359
COOLING TOWER FANS	kWe	2,345	2,328
SCRUBBER PUMPS	kWe	74	74
DESULFURIZER UNIT	kWe	5,662	5,703
GAS TURBINE AUXILIARIES	kWe	1,000	1,000
STEAM TURBINE AUXILIARIES	kWe	110	108
H2SO4 UNIT	kWe	(4,407)	(4,407)
MISCELLANEOUS BALANCE OF PLANT	kWe	3,190	3,193
TRANSFORMER LOSSES	kWe	2,701	2,728
NET PLANT EFFICIENCY, HHV	%	34.19	33.85
NET HEAT RATE	kJ/KWH	10,529	10,636
	BTU/kWH	9,980	10,081
CONDENSER COOLING DUTYNET POWER	10^6 kJ/H	1,366	1,335
	10^6 BTU/H	1,295	1,266
CONSUMABLES			
AS-RECEIVED COAL FEED	kg/H	230,379	230,595
	LB/H	507,986	508,461
THERMAL INPUT	kWt HHV	1,736,351	1,737,974
RAW WATER USAGE	M^3/MIN	21.64	22.19
	GPM	5,717	5,862
CARBON CAPTURED	%	90	90

Table C-9
Overall Plant Cost Summary
Phase 2 Cold Gas Cleanup Case
General Electric Gasifier – Fully Loaded GTs

UNIT	2007 Installed Cost ($1000)
ASU	232,935
Fuel receiving, preparation & feeding	95,487
Gasifier, syngas cooler & aux	250,110
Gasification foundations	16,253
Ash handling systems	46,038
Soot Recovery + SARU	5,857
Shift reactor	18,232
Syngas scrubber + LTGC	21,814
Blowback gas systems	2,161
Fuel gas piping	1,402
Gas cleanup foundations	1,543
Hg Removal	3,435
Selexol	211,949
Claus + TG Recycle	33,795
CO2 compression, dehydration + pumping	24,739
Syngas Expander	7,955
Gas turbine + generator + auxiliaries	132,256
HRSG, ducting + stack	55,785
Steam turbine + generator + auxiliaries	53,011
Surface condenser	7,840
Feedwater system	17,590
Water makeup + pretreating	1,910
Other feedwater subsystems	3,168
Service water systems	5,979
Other boiler plant systems	6,461
Fuel oil system & nat gas	1,930
Waste water treatment	2,303
Misc. power plant equipment	2,588
Cooling water system	38,139
Accessory electric plant	87,881
Instrumentation & controls	28,121
Improvement to site	20,196
Buildings & structures	18,684
Total	1,457,549
Total $/kW	**2,703**

Table C-10
Overall Plant Cost Summary
Phase 2 Warm Gas Cleanup Cases
General Electric Gasifier – Fully Loaded GTs

UNIT	2007 Installed Cost ($1000)
Syngas Temperature	**215°C**
ASU	239,884
Fuel receiving, preparation & feeding	98,335
Gasifier, syngas cooler & aux	257,571
Gasification foundations	16,738
Ash handling systems	47,412
Soot Recovery + SARU	6,032
Warm gas desulfurization	30,567
H2SO4 unit	54,837
Shift reactor	18,904
Syngas scrubber	22,618
Blowback gas systems	2,225
Fuel gas piping	2,149
Gas cleanup foundations	1,600
Hg Removal, CO2 separation	137,696
CO2 purification / refrigeration + pumping	113,460
Syngas Expander	11,948
Gas turbine + generator + auxiliaries	132,256
HRSG, ducting + stack	57,761
Steam turbine + generator + auxiliaries	54,284
Surface condenser	7,284
Feedwater system	28,333
Water makeup + pretreating	1,938
Other feedwater subsystems	3,244
Service water systems	6,157
Other boiler plant systems	6,617
Fuel oil system & nat gas	1,930
Waste water treatment	2,372
Misc. power plant equipment	2,622
Cooling water system	35,746
Accessory electric plant	89,045
Instrumentation & controls	28,960
Improvement to site	20,799
Buildings & structures	19,241
Total	1,560,562
Total $/kW	**2,628**

Table C-11
Cost of Electricity
Phase 2 Cold Gas Cleanup Case
General Electric Gasifier – Fully Loaded GTs
2007 $

Net power, MW	539.29
Capacity factor (CF), %	80
Total plant cost (TPC), $	1,457,548,956
6 month labor cost	13,181,137
1 month maintenance materials	2,286,810
1 month non-fuel consumables	511,559
1 month waste disposal	296,347
25% of 1 month fuel cost at 100% CF	1,697,010
2% of TPC	29,150,979
60 day supply of fuel & consumables at 100% CF	14,399,210
0.5% of TPC (spare parts)	7,287,745
Initial catalyst & chemicals cost, $	7,217,403
Land	900,000
Other owners's costs (15% of TPC)	218,632,343
Financing costs	39,353,822
Total overnight cost (TOC), $	1,792,463,322
Fixed operating cost for initial year of operation (OCF), $	55,513,252
Annual feed cost at above CF for initial year (OCV1), $	65,165,186
Other annual variable operating cost at above CF for initial year (OCV2), $	29,709,280
Annual CO2 transporting, storing, and monitoring cost at above CF for initial year (OCV3), $	19,843,753
Annual byproduct revenues at above CF for initial year (OCV4), $	3,830,926
1st year cost of electricity (COE) w/o CO2 TS&M, $/MWh	**97.7**
1st year cost of electricity (COE), $/MWh	**103.0**

Table C-12
Cost of Electricity
Phase 2 Warm Gas Cleanup Cases
General Electric Gasifier – Fully Loaded GTs
2007 $

Syngas Temperature	**215°C**
Net power, MW	593.86
Capacity factor (CF), %	80
Total plant cost (TPC), $	1,560,562,198
6 month labor cost	13,725,934
1 month maintenance materials	2,421,712
1 month non-fuel consumables	900,799
1 month waste disposal	378,973
25% of 1 month fuel cost at 100% CF	1,769,785
2% of TPC	31,211,244
60 day supply of fuel & consumables at 100% CF	15,741,249
0.5% of TPC (spare parts)	7,802,811
Initial catalyst & chemicals cost, $	12,948,344
Land	900,000
Other owners's costs	234,084,330
Financing costs	42,135,179
Total overnight cost (TOC), $	1,924,582,558
Fixed operating cost for initial year of operation (OCF), $	58,663,113
Annual feed cost at above CF for initial year (OCV1), $	67,959,745
Other annual variable operating cost at above CF for initial year (OCV2), $	35,534,247
Annual CO2 transporting, storing, and monitoring cost at above CF for initial year (OCV3), $	20,694,737
Annual byproduct revenues at above CF for initial year (OCV4), $	12,288,311
1st year cost of electricity (COE) w/o CO2 TS&M, $/MWh	**93.5**
1st year cost of electricity (COE), $/MWh	**98.5**

Table C-13
Phase 2 Avoided CO_2 Costs (2007 $)

Gasifier	**Phillips 66**		**GE**	
Syngas Cleanup	**Cold Gas**	**Warm Gas**	**Cold Gas**	**Warm Gas**
Decarbonization Technology	**Selexol**	**TDA PSA**	**Selexol**	**TDA PSA**
Net Power, MW	515.25	572.40	539.29	593.86
CO2 Emitted, ST/h	55.2528	57.8866	56.4400	58.7559
CO2 Emitted, ST/MWh	0.1072	0.1011	0.1047	0.0989
COE, $/MWh	105.15	96.58	102.98	98.46
Reference IGCC Case				
Avoided Cost (1st Year), $/ST	44.90	33.17	38.53	32.32
Avoided Cost (1st Year), $/tonne	49.50	36.57	42.48	35.63
Reference SCPC Case				
Avoided Cost (1st Year), $/ST	60.82	49.40	57.83	51.66
Avoided Cost (1st Year), $/tonne	67.05	54.46	63.76	56.95
Reference IGCC w/o CO2 capture				
Net Power, MW	625.06		622.05	
CO2 Emitted, ST/MWh	0.8550		0.8615	
COE, $/MWh (See Notes 1 & 2)	71.58		73.82	
Reference SCPC w/o CO2 capture				
Net Power, MW	549.99			
CO2 Emitted, ST/MWh	0.8840			
COE, $/MWh (See Note 1)	57.91			

Notes

1. Variable cost corrected for capacity factor
2. Byproduct sulfur credit taken since PSA cases take credit for H2SO4 produced.

Appendix D
Conoco Phillips Gasifier Equipped with SelexolTM Based Cold Gas Cleanup System – Case 4 from DOE/NETL-2010/1397

Pages from the Original Report provided here for reference.

NATIONAL ENERGY TECHNOLOGY LABORATORY

Cost and Performance Baseline for Fossil Energy Plants

Volume 1: Bituminous Coal and Natural Gas to Electricity

Revision 2, November 2010

DOE/NETL-2010/1397

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3.3.8 Case 4 - E-Gas™ IGCC Power Plant with CO_2 Capture

This case is configured to produce electric power with CO_2 capture. The plant configuration is the same as Case 3, namely two gasifier trains, two advanced F class turbines, two HRSGs, and one steam turbine. The gross power output from the plant is constrained by the capacity of the two CTs, and since the CO_2 capture and compression process increases the auxiliary load on the plant, the net output is significantly reduced relative to Case 3.

The process description for Case 4 is similar to Case 3 with several notable exceptions to accommodate CO_2 capture. A BFD and stream tables for Case 4 are shown in Exhibit 3-58 and Exhibit 3-59, respectively. Instead of repeating the entire process description, only differences from Case 3 are reported here.

Coal Preparation and Feed Systems

No differences from Case 3.

Gasification

The gasification process is the same as Case 3 with the exception that total coal feed to the two gasifiers is 5,271 tonnes/day (5,811 TPD) (stream 8) and the ASU provides 4,234 tonnes/day (4,668 TPD) of 95 mol% oxygen to the gasifier and Claus plant (streams 5 and 3).

Raw Gas Cooling/Particulate Removal

Raw gas cooling and particulate removal are the same as Case 3 with the exception that approximately 418,710 kg/hr (923,082 lb/hr) of saturated steam at 13.8 MPa (2,000 psia) is generated in the SGC.

Syngas Scrubber/Sour Water Stripper

No differences from Case 3.

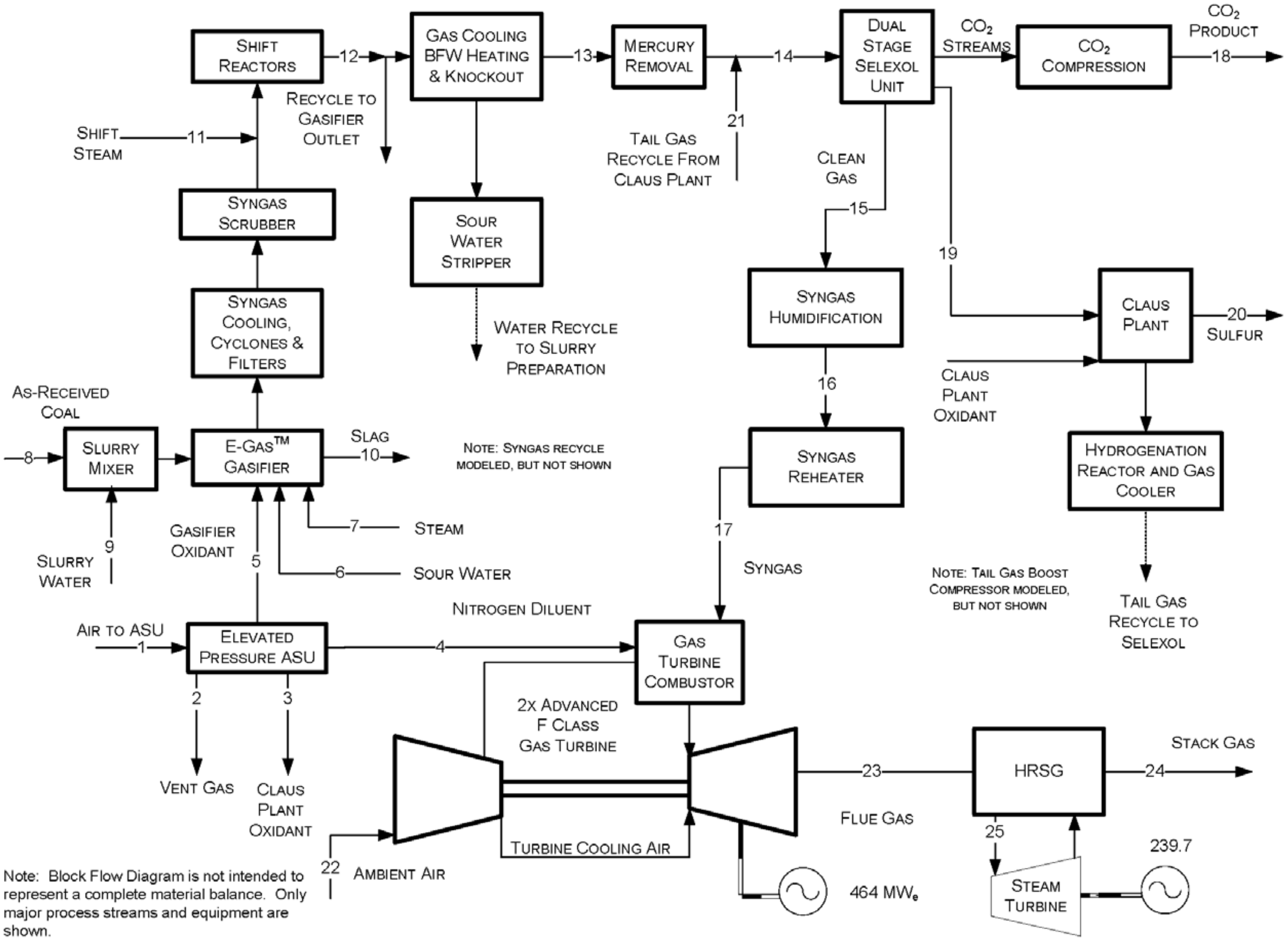
Sour Gas Shift (SGS)

The SGS process was described in Section 3.1.3. In Case 4 steam (stream 11) is added to the syngas exiting the scrubber to adjust the H_2O:CO molar ratio to approximately 2.25:1 prior to the first WGS reactor. The hot syngas exiting the first stage of SGS is used to preheat a portion of the water used to humidify the clean syngas leaving the AGR. The final stage of SGS brings the overall conversion of the CO to CO_2 to 98.5 percent. The syngas exiting the final stage of SGS still contains 1.2 vol% CH_4, which is subsequently oxidized to CO_2 in the CT and results in a carbon capture of 90.4 percent. The warm syngas exiting the second stage of the SGS at 204°C (400°F) (stream 12) is cooled to 201°C (393°F) by preheating the syngas entering the first SGS reactor. The SGS catalyst also serves to hydrolyze COS thus eliminating the need for a separate COS hydrolysis reactor. The syngas is further dehydrated and cooled to 35°C (95°F) in syngas coolers prior to the mercury removal beds.

Mercury Removal and Acid Gas Removal

Mercury removal is the same as in Case 3.

Exhibit 3-58 Case 4 Block Flow Diagram, E-Gas™ IGCC with CO_2 Capture



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown.

Exhibit 3-59 Case 4 Stream Table, E-Gas™ IGCC with CO_2 Capture

	1	2	3	4	5	6	7	8	9	10	11	12	13
V-L Mole Fraction													
Ar	0.0092	0.0209	0.0318	0.0023	0.0318	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0054	0.0071
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0125	0.0164
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0028	0.0037
CO_2	0.0003	0.0071	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3103	0.4090
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.4128	0.5441
H_2O	0.0099	0.1780	0.0000	0.0003	0.0000	1.0000	1.0000	0.0000	0.9963	0.0000	1.0000	0.2376	0.0015
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0048	0.0063
N_2	0.7732	0.6187	0.0178	0.9919	0.0178	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0090	0.0119
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0037	0.0000	0.0000	0.0046	0.0000
O_2	0.2074	0.1754	0.9504	0.0054	0.9504	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	26,685	1,231	145	19,704	5,338	1,287	4,969	0	5,009	0	9,357	37,866	26,948
V-L Flowrate (kg/hr)	770,042	33,603	4,654	552,893	171,782	23,193	89,523	0	90,226	0	168,566	748,369	547,649
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	219,635	0	22,418	0	0	0
Temperature (°C)	15	19	32	93	32	154	343	15	171	1,038	288	204	35
Pressure (MPa, abs)	0.10	0.11	0.86	2.65	0.86	5.79	5.10	0.10	5.79	4.24	5.52	4.07	3.79
Enthalpy (kJ/kg)^A	30.23	36.49	26.67	92.50	26.67	599.34	3,063.97	---	673.50	---	2,918.18	873.73	40.91
Density (kg/m^3)	1.2	1.5	11.0	24.4	11.0	857.7	20.1	---	836.0	---	25.6	20.6	30.9
V-L Molecular Weight	28.857	27.295	32.181	28.060	32.181	18.015	18.015	---	18.012	---	18.015	19.764	20.322
V-L Flowrate (lb_{mol}/hr)	58,830	2,714	319	43,440	11,768	2,838	10,955	0	11,044	0	20,628	83,479	59,411
V-L Flowrate (lb/hr)	1,697,652	74,082	10,260	1,218,920	378,715	51,133	197,365	0	198,914	0	371,625	1,649,872	1,207,359
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	484,212	0	49,422	0	0	0
Temperature (°F)	59	67	90	199	90	310	650	59	340	1,900	550	400	95
Pressure (psia)	14.7	16.4	125.0	384.0	125.0	840.0	740.0	14.7	840.0	615.0	800.0	590.0	550.0
Enthalpy (Btu/lb)^A	13.0	15.7	11.5	39.8	11.5	257.7	1,317.3	---	289.6	---	1,254.6	375.6	17.6
Density (lb/ft^3)	0.076	0.095	0.687	1.521	0.687	53.543	1.257	---	52.192	---	1.597	1.285	1.928
	A - Reference conditions are 32.02 F & 0.089 PSIA												

Exhibit 3-59 Case 4 Stream Table, E-Gas™ IGCC with CO_2 Capture (Continued)

	14	15	16	17	18	19	20	21	22	23	24	25
V-L Mole Fraction												
Ar	0.0071	0.0119	0.0114	0.0114	0.0002	0.0011	0.0000	0.0068	0.0092	0.0090	0.0090	0.0000
CH_4	0.0159	0.0262	0.0251	0.0251	0.0008	0.0045	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0038	0.0063	0.0060	0.0060	0.0001	0.0007	0.0000	0.0076	0.0000	0.0000	0.0000	0.0000
CO_2	0.4172	0.0352	0.0338	0.0338	0.9945	0.7021	0.0000	0.6907	0.0003	0.0082	0.0082	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.5341	0.8964	0.8588	0.8588	0.0044	0.0591	0.0000	0.2007	0.0000	0.0000	0.0000	0.0000
H_2O	0.0015	0.0001	0.0421	0.0421	0.0000	0.0162	0.0000	0.0017	0.0099	0.1246	0.1246	1.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0062	0.0000	0.0000	0.0000	0.0000	0.2153	0.0000	0.0026	0.0000	0.0000	0.0000	0.0000
N_2	0.0142	0.0239	0.0229	0.0229	0.0001	0.0010	0.0000	0.0901	0.7732	0.7529	0.7529	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2074	0.1052	0.1052	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	27,761	16,435	17,155	17,155	10,499	800	0	812	110,253	139,694	139,694	34,515
V-L Flowrate (kg/hr)	575,208	83,813	96,780	96,780	459,855	31,069	0	27,560	3,181,557	3,831,230	3,831,230	621,792
Solids Flowrate (kg/hr)	0	0	0	0	0	0	5,494	0	0	0	0	0
Temperature (°C)	34	34	108	193	51	48	176	38	15	562	132	534
Pressure (MPa, abs)	3.8	3.757	3.206	3.172	15.270	0.163	0.119	5.512	0.101	0.105	0.105	12.512
Enthalpy (kJ/kg)^A	39.0	196.106	903.063	1,360.503	-162.349	62.132	---	1.774	30.227	839.766	348.188	3,432.696
Density (kg/m^3)	31.3	7.4	5.6	4.6	641.8	2.4	5,283.7	83.6	1.2	0.4	0.9	36.7
V-L Molecular Weight	21	5.100	5.642	5.642	43.800	38.814	---	33.925	28.857	27.426	27.426	18.015
V-L Flowrate (lb_{mol}/hr)	61,202	36,233	37,820	37,820	23,146	1,765	0	1,791	243,066	307,972	307,972	76,092
V-L Flowrate (lb/hr)	1,268,117	184,776	213,363	213,363	1,013,807	68,496	0	60,759	7,014,133	8,446,417	8,446,417	1,370,817
Solids Flowrate (lb/hr)	0	0	0	0	0	0	12,112	0	0	0	0	0
Temperature (°F)	94	94	227	380	124	119	349	100	59	1,044	270	994
Pressure (psia)	545.0	545.0	465.0	460.0	2,214.7	23.7	17.3	799.5	14.7	15.2	15.2	1,814.7
Enthalpy (Btu/lb)^A	16.8	84.3	388.2	584.9	-69.8	26.7	---	0.8	13.0	361.0	149.7	1,475.8
Density (lb/ft^3)	2	0.461	0.352	0.285	40.067	0.149	329.851	5.220	0.076	0.026	0.053	2.293

The AGR process in Case 4 is a two stage Selexol process where H_2S is removed in the first stage and CO_2 in the second stage of absorption as previously described in Section 3.1.5. The process results in three product streams, the clean syngas, a CO_2-rich stream, and an acid gas feed to the Claus plant. The acid gas (stream 19) contains 21.5 percent H_2S and 70 percent CO_2 with the balance primarily H_2. The CO_2-rich stream is discussed further in the CO_2 compression section.

CO_2 Compression and Dehydration

CO_2 from the AGR process is flashed at three pressure levels to separate CO_2 and decrease H_2 losses to the CO_2 product pipeline. The HP CO_2 stream is flashed at 2.0 MPa (289.7 psia), compressed, and recycled back to the CO_2 absorber. The MP CO_2 stream is flashed at 1.0 MPa (149.7 psia). The LP CO_2 stream is flashed at 0.1 MPa (16.7 psia), compressed to 1.0 MPa (149.5 psia), and combined with the MP CO_2 stream. The combined stream is compressed from 1.0 MPa (149.5 psia) to a SC condition at 15.3 MPa (2215 psia) using a multiple-stage, intercooled compressor. During compression, the CO_2 stream is dehydrated to a dewpoint of -40ºC (-40°F) with triethylene glycol. The raw CO_2 stream from the Selexol process contains over 99 percent CO_2. The CO_2 (stream 18) is transported to the plant fence line and is sequestration ready. CO_2 TS&M costs were estimated using the methodology described in Section 2.7.

Claus Unit

The Claus plant is the same as Case 3 with the following exceptions:

- 5,494 kg/hr (12,112 lb/hr) of sulfur (stream 20) are produced
- The waste heat boiler generates 12,679 kg/hr (27,953 lb/hr) of 3.0 MPa (430 psia) steam, which provides all of the Claus plant process needs and provides some additional steam to the medium pressure steam header.

Power Block

Clean syngas from the AGR plant is partially humidified to 4 percent because the nitrogen available from the ASU is insufficient to provide adequate dilution. The moisturized syngas is reheated (stream 17) to 193°C (380°F) using HP BFW, diluted with nitrogen (stream 4), and then enters the CT burner. There is no integration between the CT and the ASU in this case. The exhaust gas (stream 23) exits the CT at 562°C (1044°F) and enters the HRSG where additional heat is recovered. The FG exits the HRSG at 132°C (270°F) (stream 24) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced commercially available steam turbine using a 12.4 MPa/534°C/534°C (1800 psig/994°F/994°F) steam cycle.

Air Separation Unit (ASU)

The elevated pressure ASU is the same as in other cases and produces 4,234 tonnes/day (4,668 TPD) of 95 mol% oxygen and 14,230 tonnes/day (15,686 TPD) of nitrogen. There is no integration between the ASU and the CT.

3.3.9 Case 4 Performance Results

The Case 4 modeling assumptions were presented previously in Section 3.3.3.

The plant produces a net output of 514 MWe at a net plant efficiency of 31.0 percent (HHV basis). Overall performance for the entire plant is summarized in Exhibit 3-60, which includes auxiliary power requirements. The ASU accounts for nearly 58 percent of the auxiliary load between the main air compressor, the nitrogen compressor, the oxygen compressor, and ASU auxiliaries. The two-stage Selexol process and CO_2 compression account for an additional 27 percent of the auxiliary power load. The BFW pumps and cooling water system (CWPs and cooling tower fan) comprise nearly 6 percent of the load, leaving 9 percent of the auxiliary load for all other systems.

Exhibit 3-60 Case 4 Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Gas Turbine Power	464,000
Sweet Gas Expander Power	0
Steam Turbine Power	239,700
TOTAL POWER, kWe	**703,700**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	470
Coal Milling	2,260
Sour Water Recycle Slurry Pump	200
Slag Handling	1,160
Air Separation Unit Auxiliaries	1,000
Air Separation Unit Main Air Compressor	65,670
Oxygen Compressor	9,010
Nitrogen Compressors	35,340
CO_2 Compressor	31,380
Boiler Feedwater Pumps	4,160
Condensate Pump	310
Syngas Recycle Compressor	520
Circulating Water Pump	4,670
Ground Water Pumps	520
Cooling Tower Fans	2,410
Scrubber Pumps	400
Acid Gas Removal	19,900
Gas Turbine Auxiliaries	1,000
Steam Turbine Auxiliaries	100
Claus Plant/TGTU Auxiliaries	250
Claus Plant TG Recycle Compressor	3,700
Miscellaneous Balance of Plant[2]	3,000
Transformer Losses	2,660
TOTAL AUXILIARIES, kWe	**190,090**
NET POWER, kWe	**513,610**
Net Plant Efficiency, % (HHV)	31.0
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	11,604 (10,998)
CONDENSER COOLING DUTY 10^6 kJ/hr (10^6 Btu/hr)	**1,403 (1,330)**
CONSUMABLES	
As-Received Coal Feed, kg/hr (lb/hr)	219,635 (484,212)
Thermal Input[1], kWt	1,655,503
Raw Water Withdrawal, m^3/min (gpm)	21.6 (5,717)
Raw Water Consumption, m^3/min (gpm)	17.5 (4,631)

1 - HHV of As-Received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)

2 - Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

Environmental Performance

The environmental targets for emissions of Hg, NOx, SO_2, CO_2, and PM were presented in Section 2.4. A summary of the plant air emissions for Case 4 is presented in Exhibit 3-61.

Exhibit 3-61 Case 4 Air Emissions

	kg/GJ ($lb/10^6$ Btu)	Tonne/year (tons/year) 80% CF	kg/MWh (lb/MWh)
SO_2	0.001 (0.002)	39 (43)	0.008 (.02)
NOx	0.021 (0.049)	885 (976)	0.180 (.396)
Particulates	0.003 (0.0071)	127 (141)	0.026 (.057)
Hg	2.46E-7 (5.71E-7)	0.010 (0.011)	2.08E-6 (4.59E-6)
CO_2	8.5 (19.7)	354,267 (390,512)	72 (158)
CO_2[1]			98 (217)

[1] CO_2 emissions based on net power instead of gross power

The low level of SO_2 emissions is achieved by capture of the sulfur in the gas by the two-stage Selexol AGR process. The CO_2 capture target results in the sulfur compounds being removed to a greater extent than required in the environmental targets of Section 2.4. The clean syngas exiting the AGR process has a sulfur concentration of approximately 5 ppmv. This results in a concentration in the FG of less than 1 ppmv. The H_2S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is hydrogenated to convert all sulfur species to H_2S, and then recycled back to the Selexol, thereby eliminating the need for a tail gas treatment unit.

NOx emissions are limited by the use of humidification and nitrogen dilution to 15 ppmvd (NO_2 @ 15 percent O_2). Ammonia in the syngas is removed with process condensate prior to the low-temperature AGR process and ultimately destroyed in the Claus plant burner. This helps lower NOx levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a cyclone and a barrier filter in addition to the syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Ninety five percent of mercury is captured from the syngas by an activated carbon bed.

Ninety five percent of the CO_2 from the syngas is captured in the AGR system and compressed for sequestration. The overall carbon removal is 90.4 percent.

The carbon balance for the plant is shown in Exhibit 3-62. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not neglected in the carbon balance below since the Aspen model accounts for air components throughout. Carbon leaves the plant as unburned carbon in the slag, CO_2 in the stack gas, ASU vent gas and the captured CO_2 product. The carbon capture efficiency is defined as the amount of carbon in the

CO_2 product stream relative to the amount of carbon in the coal less carbon contained in the slag, represented by the following fraction:

(Carbon in Product for Sequestration)/[(Carbon in the Coal)-(Carbon in Slag)] or
276,728/(308,659-2,469) *100 or
90.4 percent

In revision 1 of this report, the reported CO_2 capture efficiency was 88.4 percent. The high methane content of the syngas, relative to the GEE and Shell cases, prevented reaching the nominal 90 percent CO_2 capture. In order to achieve 90 percent capture, the two-stage Selexol CO_2 removal efficiency was increased from 92 to 95 percent.

Exhibit 3-62 Case 4 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Coal	140,006 (308,659)	**Slag**	1,120 (2,469)
Air (CO_2)	537 (1,185)	**Stack Gas**	13,796 (30,416)
		ASU Vent	105 (231)
		CO_2 Product	125,522 (276,728)
Total	140,543 (309,844)	**Total**	140,543 (309,844)

Exhibit 3-63 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant, sulfur emitted in the stack gas, and sulfur in the CO_2 product. Sulfur in the slag is considered to be negligible.

Exhibit 3-63 Case 4 Sulfur Balance

Sulfur In, kg/hr (lb/hr)		Sulfur Out, kg/hr (lb/hr)	
Coal	5,505 (12,136)	**Elemental Sulfur**	5,494 (12,112)
		Stack Gas	3 (6)
		CO_2 Product	8 (18)
Total	5,505 (12,136)	**Total**	5,505 (12,136)

Exhibit 3-64 shows the overall water balance for the plant. The exhibit is presented in an identical manner for Cases 1 through 3.

Exhibit 3-64 Case 4 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Slag Handling	0.49 (128)	0.49 (128)	0.0 (0)	0.0 (0)	0.0 (0)
Slurry Water	1.51 (398)	1.51 (398)	0.0 (0)	0.0 (0)	0.0 (0)
Quench/Wash	0.0 (0)	0.0 (0)	0.0 (0)	0.0 (0)	0.0 (0)
Humidifier	0.2 (61)	0.2 (61)	0.0 (0)	0.0 (0)	0.0 (0)
SWS Blowdown	0.0 (0)	0.0 (0)	0.0 (0)	0.03 (7)	-0.03 (-7)
Condenser Makeup	**4.5 (1,193)**	**0.0 (0)**	**4.5 (1,193)**	**0.0 (0)**	**4.5 (1,193)**
Gasifier Steam	*1.5 (395)*		*1.5 (395)*		
Shift Steam	*2.8 (743)*		*2.8 (743)*		
GT Steam Dilution	*0.21 (55)*		*0.21 (55)*		
BFW Makeup					
Cooling Tower	**18.2 (4,798)**	**1.0 (274)**	**17.1 (4,524)**	**4.1 (1,079)**	**13.0 (3,445)**
BFW Blowdown		*0.21 (55)*	*-0.21 (-55)*		
SWS Blowdown		*0.26 (68)*	*-0.26 (-68)*		
SWS Excess Water		*0.6 (152)*	*-0.6 (-152)*		
Humidifier Tower Blowdown					
Total	**24.9 (6,578)**	**3.3 (861)**	**21.6 (5,717)**	**4.1 (1,086)**	**17.5 (4,631)**

Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown for the following subsystems in Exhibit 3-65 through Exhibit 3-67:

- Coal gasification and ASU
- Syngas cleanup including sulfur recovery and tail gas recycle
- Combined cycle power generation, steam, and FW

An overall plant energy balance is provided in tabular form in Exhibit 3-68. The power out is the combined CT and steam turbine power prior to generator losses. The power at the generator terminals (shown in Exhibit 3-60) is calculated by multiplying the power out by a combined generator efficiency of 98.4 percent.

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Exhibit 3-65 Case 4 Coal Gasification and Air Separation Unit Heat and Mass Balance Schematic

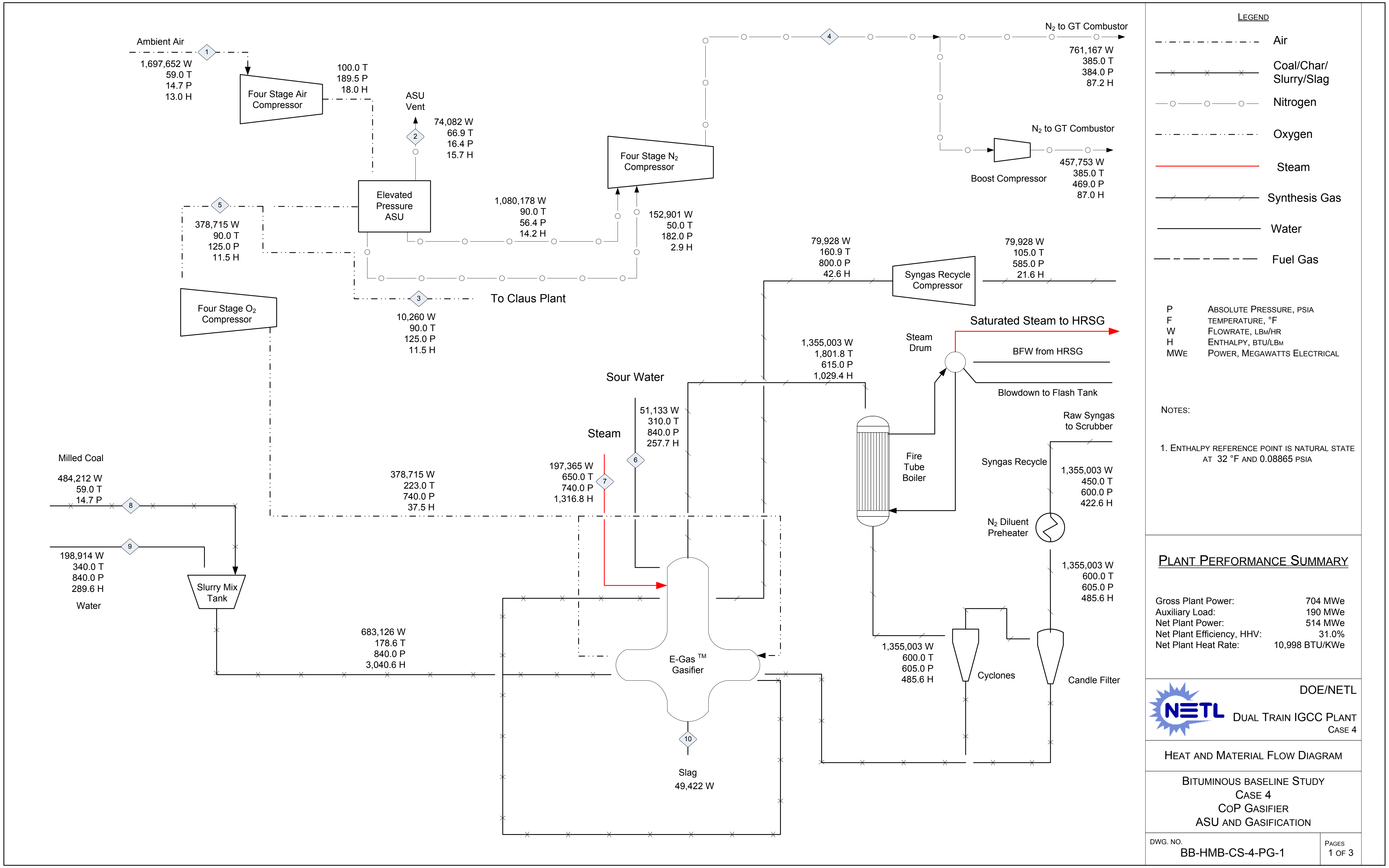

Exhibit 3-66 Case 4 Syngas Cleanup Heat and Mass Balance Schematic

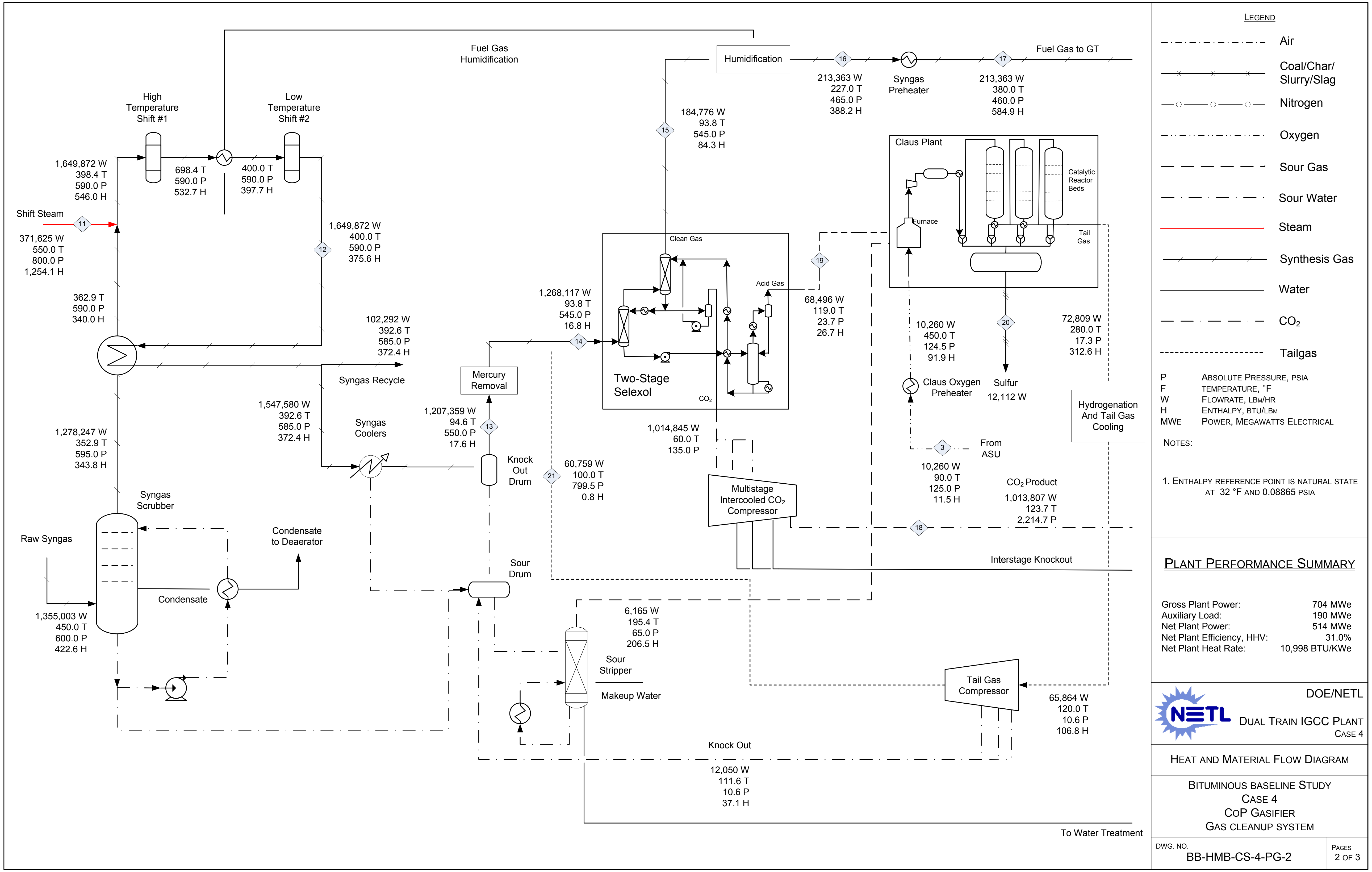

Exhibit 3-67 Case 4 Combined Cycle Power Generation Heat and Mass Balance Schematic

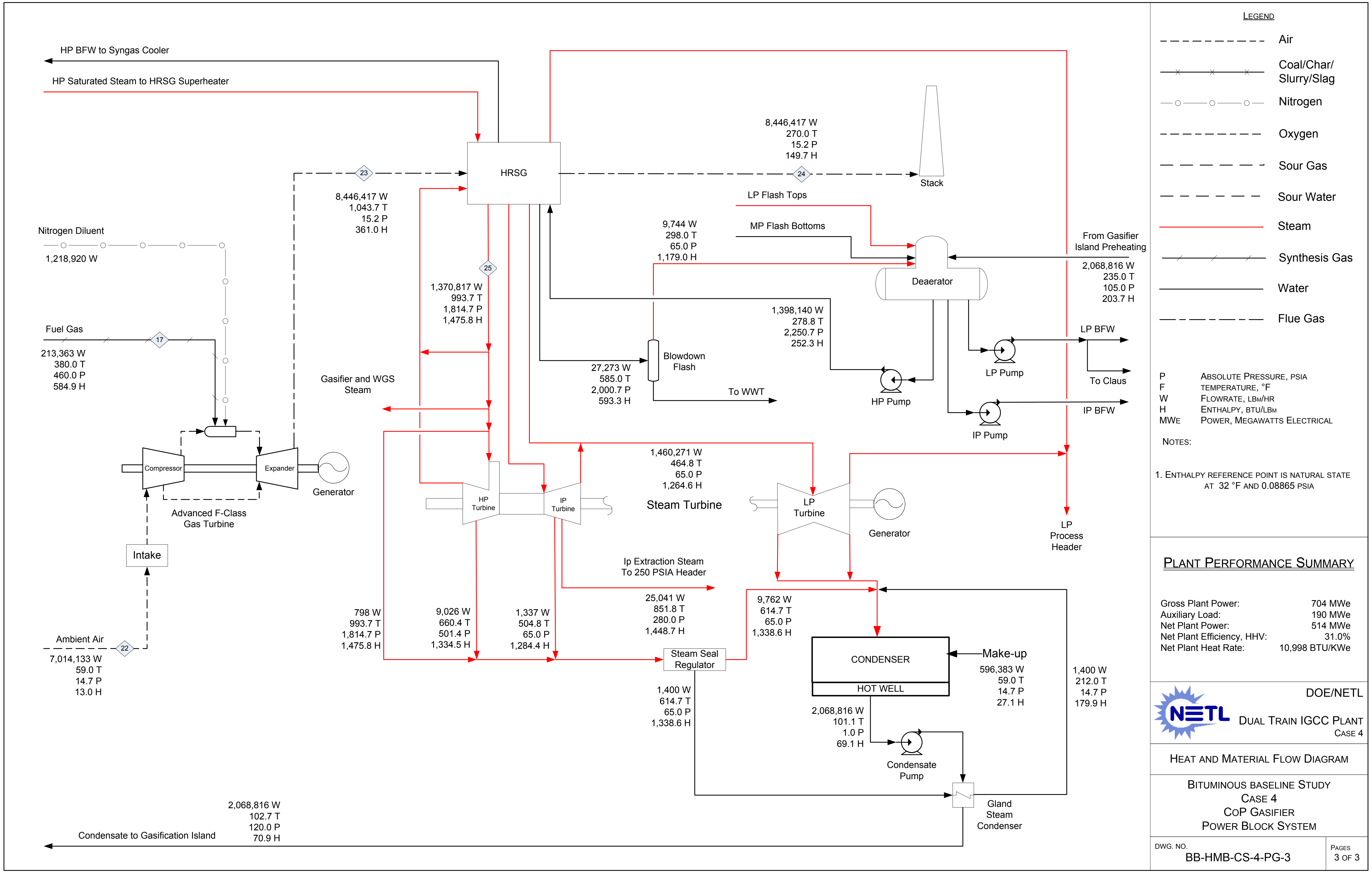

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Exhibit 3-68 Case 4 Overall Energy Balance (0°C [32°F] Reference)

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Coal	5,960 (5,649)	5.0 (4.7)		5,965 (5,654)
ASU Air		23.3 (22.1)		23 (22)
GT Air		96.2 (91.2)		96 (91)
Water		81.4 (77.1)		81 (77)
Auxiliary Power			684 (649)	684 (649)
TOTAL	**5,960 (5,649)**	**205.8 (195.1)**	**684 (649)**	**6,850 (6,492)**
Heat Out GJ/hr (MMBtu/hr)				
ASU Vent		1.2 (1.2)		1 (1)
Slag	37 (35)	25.2 (23.9)		62 (59)
Sulfur	51 (48)	0.6 (0.6)		52 (49)
CO_2		-74.7 (-70.8)		-75 (-71)
Cooling Tower Blowdown		30.3 (28.8)		30 (29)
HRSG Flue Gas		1,334 (1,264)		1,334 (1,264)
Condenser		1,408 (1,334)		1,408 (1,334)
Non-Condenser Cooling Tower Loads*		755 (716)		755 (716)
Process Losses**		749 (710)		749 (710)
Power			2,533 (2,401)	2,533 (2,401)
TOTAL	**88 (83)**	**4,229 (4,008)**	**2,533 (2,401)**	**6,850 (6,492)**

* Includes ASU compressor intercoolers, CO_2 compressor intercoolers, sour water stripper condenser, syngas cooler (low level heat rejection) and extraction air cooler.

** Calculated by difference to close the energy balance

3.3.10 Case 4 - Major Equipment List

Major equipment items for the CoP gasifier with CO_2 capture are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 3.3.11. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

ACCOUNT 1 COAL HANDLING

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	572 tonne/hr (630 tph)	2	0
3	Conveyor No. 1	Belt	1,134 tonne/hr (1,250 tph)	1	0
4	Transfer Tower No. 1	Enclosed	N/A	1	0
5	Conveyor No. 2	Belt	1,134 tonne/hr (1,250 tph)	1	0
6	As-Received Coal Sampling System	Two-stage	N/A	1	0
7	Stacker/Reclaimer	Traveling, linear	1,134 tonne/hr (1,250 tph)	1	0
8	Reclaim Hopper	N/A	45 tonne (50 ton)	2	1
9	Feeder	Vibratory	181 tonne/hr (200 tph)	2	1
10	Conveyor No. 3	Belt w/ tripper	363 tonne/hr (400 tph)	1	0
11	Crusher Tower	N/A	N/A	1	0
12	Coal Surge Bin w/ Vent Filter	Dual outlet	181 tonne (200 ton)	2	0
13	Crusher	Impactor reduction	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2	0
14	As-Fired Coal Sampling System	Swing hammer	N/A	1	1
15	Conveyor No. 4	Belt w/tripper	363 tonne/hr (400 tph)	1	0
16	Transfer Tower No. 2	Enclosed	N/A	1	0
17	Conveyor No. 5	Belt w/ tripper	363 tonne/hr (400 tph)	1	0
18	Coal Silo w/ Vent Filter and Slide Gates	Field erected	363 tonne (400 ton)	6	0

ACCOUNT 2 COAL PREPARATION AND FEED

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Feeder	Vibratory	82 tonne/h (90 tph)	3	0
2	Conveyor No. 6	Belt w/tripper	245 tonne/h (270 tph)	1	0
3	Rod Mill Feed Hopper	Dual Outlet	481 tonne (530 ton)	1	0
4	Weigh Feeder	Belt	118 tonne/h (130 tph)	2	0
5	Rod Mill	Rotary	118 tonne/h (130 tph)	2	0
6	Slurry Water Storage Tank with Agitator	Field erected	298,179 liters (78,770 gal)	2	0
7	Slurry Water Pumps	Centrifugal	833 lpm (220 gpm)	2	1
8	Trommel Screen	Coarse	172 tonne/h (190 tph)	2	0
9	Rod Mill Discharge Tank with Agitator	Field erected	390,052 liters (103,040 gal)	2	0
10	Rod Mill Product Pumps	Centrifugal	3,407 lpm (900 gpm)	2	2
11	Slurry Storage Tank with Agitator	Field erected	1,170,080 liters (309,100 gal)	2	0
12	Slurry Recycle Pumps	Centrifugal	6,435 lpm (1,700 gpm)	2	2
13	Slurry Product Pumps	Positive displacement	3,407 lpm (900 gpm)	2	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	1,101,555 liters (291,000 gal)	2	0
2	Condensate Pumps	Vertical canned	8,669 lpm @ 91 m H2O (2,290 gpm @ 300 ft H2O)	2	1
3	Deaerator (integral w/ HRSG)	Horizontal spray type	606,907 kg/hr (1,338,000 lb/hr)	2	0
4	Intermediate Pressure Feedwater Pump	Horizontal centrifugal, single stage	8,555 lpm @ 27 m H2O (2,260 gpm @ 90 ft H2O)	2	1
5	High Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 6,057 lpm @ 1,859 m H2O (1,600 gpm @ 6,100 ft H2O)	2	1
6	High Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 1,287 lpm @ 223 m H2O (340 gpm @ 730 ft H2O)	2	1
7	Auxiliary Boiler	Shop fabricated, water tube	18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	1	0
8	Service Air Compressors	Flooded Screw	28 m3/min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
9	Instrument Air Dryers	Duplex, regenerative	28 m3/min (1,000 scfm)	2	1
10	Closed Cylce Cooling Heat Exchangers	Plate and frame	430 GJ/hr (407 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	154,066 lpm @ 21 m H2O (40,700 gpm @ 70 ft H2O)	2	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 107 m H2O (1,000 gpm @ 350 ft H2O)	1	1
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 76 m H2O (700 gpm @ 250 ft H2O)	1	1
14	Raw Water Pumps	Stainless steel, single suction	5,602 lpm @ 18 m H2O (1,480 gpm @ 60 ft H2O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	2,801 lpm @ 268 m H2O (740 gpm @ 880 ft H2O)	4	1
16	Filtered Water Pumps	Stainless steel, single suction	3,066 lpm @ 49 m H2O (810 gpm @ 160 ft H2O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	1,472,525 liter (389,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	1,855 lpm (490 gpm)	2	0
19	Liquid Waste Treatment System		10 years, 24-hour storm	1	0

ACCOUNT 4 GASIFIER, ASU AND ACCESSORIES INCLUDING LOW TEMPERATURE HEAT RECOVERY AND FUEL GAS SATURATION

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized two-stage, slurry-feed entrained bed	2,903 tonne/day, 4.2 MPa (3,200 tpd, 614.96 psia)	2	0
2	Synthesis Gas Cooler	Fire-tube boiler	337,926 kg/hr (745,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	337,926 kg/hr (745,000 lb/hr) Design efficiency 90%	2	0
4	Candle Filter	Pressurized filter with pulse-jet cleaning	metallic filters	2	0
5	Syngas Scrubber Including Sour Water Stripper	Vertical upflow	337,926 kg/hr (745,000 lb/hr)	2	0
6	Raw Gas Coolers	Shell and tube with condensate drain	386,007 kg/hr (851,000 lb/hr)	8	0
7	Raw Gas Knockout Drum	Vertical with mist eliminator	303,000 kg/hr, 35°C, 3.8 MPa (668,000 lb/hr, 95°F, 555 psia)	2	0
8	Saturation Water Economizers	Shell and tube	36 GJ/hr (34 MMBtu/hr)	2	0
9	Fuel Gas Saturator	Vertical tray tower	53,070 kg/hr, 108°C, 3.8 MPa (117,000 lb/hr, 227°F, 545 psia)	2	0
10	Saturator Water Pump	Centrifugal	757 lpm @ 12 m H2O (200 gpm @ 40 ft H2O)	2	2
11	Synthesis Gas Reheater	Shell and tube	53,070 kg/hr (117,000 lb/hr)	2	0
12	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	337,926 kg/hr (745,000 lb/hr) syngas	2	0
13	ASU Main Air Compressor	Centrifugal, multi-stage	5,805 m3/min @ 1.3 MPa (205,000 scfm @ 190 psia)	2	0
14	Cold Box	Vendor design	2,359 tonne/day (2,600 tpd) of 95% purity oxygen	2	0
15	Oxygen Compressor	Centrifugal, multi-stage	1,161 m3/min (41,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia)	2	0
16	Primary Nitrogen Compressor	Centrifugal, multi-stage	3,794 m3/min (134,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)	2	0
17	Secondary Nitrogen Compressor	Centrifugal, single-stage	538 m3/min (19,000 scfm) Suction - 1.2 MPa (180 psia) Discharge - 2.7 MPa (390 psia)	2	0
18	Gasifier Purge Nitrogen Boost Compressor	Centrifugal, single-stage	1,614 m3/min (57,000 scfm) Suction - 2.6 MPa (380 psia) Discharge - 3.2 MPa (470 psia)	2	0

ACCOUNT 5A SOUR GAS SHIFT AND SYNGAS CLEANUP

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber	Sulfated carbon bed	301,185 kg/hr (664,000 lb/hr) 35°C (95°F) 3.8 MPa (550 psia)	2	0
2	Sulfur Plant	Claus type	145 tonne/day (160 tpd)	1	0
3	Water Gas Shift Reactors	Fixed bed, catalytic	411,408 kg/hr (907,000 lb/hr) 204°C (400°F) 4.1 MPa (590 psia)	4	0
4	Shift Reactor Heat Recovery Exchangers	Shell and Tube	Exchanger 1: 94 GJ/hr (89 MMBtu/hr) Exchanger 2: 3 GJ/hr (3 MMBtu/hr)	4	0
5	Acid Gas Removal Plant	Two-stage Selexol	316,154 kg/hr (697,000 lb/hr) 34°C (94°F) 3.8 MPa (545 psia)	2	0
6	Hydrogenation Reactor	Fixed bed, catalytic	36,328 kg/hr (80,090 lb/hr) 232°C (450°F) 0.1 MPa (12.3 psia)	1	0
7	Tail Gas Recycle Compressor	Centrifugal	30,316 kg/hr (66,835 lb/hr)	1	0

ACCOUNT 5B CO_2 COMPRESSION

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CO2 Compressor	Integrally geared, multi-stage centrifugal	1,141 m3/min @ 15.3 MPa (40,300 scfm @ 2,215 psia)	4	0

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gas Turbine	Advanced F class	232 MW	2	0
2	Gas Turbine Generator	TEWAC	260 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	2	0

ACCOUNT 7 HRSG, DUCTING, AND STACK

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 8.5 m (28 ft) diameter	1	0
2	Heat Recovery Steam Generator	Drum, multi-pressure with economizer section and integral deaerator	Main steam - 341,986 kg/hr, 12.4 MPa/534°C (753,949 lb/hr, 1,800 psig/994°F) Reheat steam - 298,222 kg/hr, 3.1 MPa/534°C (657,466 lb/hr, 452 psig/994°F)	2	0

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	252 MW 12.4 MPa/534°C/534°C (1800 psig/ 994°F/994°F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitiation	280 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Steam Bypass	One per HRSG	50% steam flow @ design steam conditions	2	0
4	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1,551 GJ/hr (1,470 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	1	0

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	469,391 lpm @ 30 m (124,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 2606 GJ/hr (2470 MMBtu/hr) heat duty	1	0

ACCOUNT 10 SLAG RECOVERY AND HANDLING

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	234,696 liters (62,000 gal)	2	0
2	Slag Crusher	Roll	13 tonne/hr (14 tph)	2	0
3	Slag Depressurizer	Proprietary	13 tonne/hr (14 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	140,060 liters (37,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	64,352 liters (17,000 gal)	2	
6	Slag Conveyor	Drag chain	13 tonne/hr (14 tph)	2	0
7	Slag Separation Screen	Vibrating	13 tonne/hr (14 tph)	2	0
8	Coarse Slag Conveyor	Belt/bucket	13 tonne/hr (14 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	200,627 liters (53,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	38 lpm @ 14 m H2O (10 gpm @ 46 ft H2O)	2	2
11	Grey Water Storage Tank	Field erected	64,352 liters (17,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	227 lpm @ 433 m H2O (60 gpm @ 1,420 ft H2O)	2	2
13	Slag Storage Bin	Vertical, field erected	907 tonne (1,000 tons)	2	0
14	Unloading Equipment	Telescoping chute	100 tonne/hr (110 tph)	1	0

ACCOUNT 11 ACCESSORY ELECTRIC PLANT

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CTG Step-up Transformer	Oil-filled	24 kV/345 kV, 260 MVA, 3-ph, 60 Hz	2	0
2	STG Step-up Transformer	Oil-filled	24 kV/345 kV, 280 MVA, 3-ph, 60 Hz	1	0
3	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 79 MVA, 3-ph, 60 Hz	2	0
4	Medium Voltage Auxiliary Transformer	Oil-filled	24 kV/4.16 kV, 51 MVA, 3-ph, 60 Hz	1	1
5	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 8 MVA, 3-ph, 60 Hz	1	1
6	CTG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	2	0
7	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
8	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
9	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
10	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

ACCOUNT 12 INSTRUMENTATION AND CONTROLS

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS - Processor	Microprocessor with redundant input/output	N/A	1	0
3	DCS - Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.3.11 Case 4 - Cost Estimating Results

The cost estimating methodology was described previously in Section 2.6. Exhibit 3-69 shows the total plant capital cost summary organized by cost account and Exhibit 3-70 shows a more detailed breakdown of the capital costs as well as TOC, TASC, and breakdown of owner's costs. Exhibit 3-71 shows the initial and annual O&M costs.

The estimated TOC of the CoP gasifier with CO_2 capture is $3,466/kW. Process contingency represents 3.5 percent of the TOC and project contingency represents 11.1 percent. The COE, including CO_2 TS&M costs of 5.6 mills/kWh, is 110.4 mills/kWh.

Exhibit 3-69 Case 4 Total Plant Cost Summary

Client: USDOE/NETL **Report Date:** 2010-Jan-14
Project: Bituminous Baseline Study

TOTAL PLANT COST SUMMARY

Case: Case 4 - ConocoPhillips 500MW IGCC w/ CO2
Plant Size: 513.6 MW,net **Estimate Type:** Conceptual **Cost Base (Jun)** 2007 ($x1000)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
1	COAL & SORBENT HANDLING	$14,229	$2,644	$11,035	$0	$0	**$27,908**	$2,533	$0	$6,088	**$36,529**	$71
2	COAL & SORBENT PREP & FEED	$24,241	$4,431	$14,646	$0	$0	**$43,318**	$3,889	$0	$9,441	**$56,648**	$110
3	FEEDWATER & MISC. BOP SYSTEMS	$10,074	$7,882	$10,144	$0	$0	**$28,101**	$2,651	$0	$7,106	**$37,858**	$74
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries (E-GAS)	$114,050	$0	$63,266	$0	$0	**$177,316**	$16,295	$24,521	$33,478	**$251,609**	$490
4.2	Syngas Cooling	w/4.1	$0	w/ 4.1	$0	$0	**$0**	$0	$0	$0	**$0**	$0
4.3	ASU/Oxidant Compression	$186,025	$0	w/equip.	$0	$0	**$186,025**	$18,031	$0	$20,406	**$224,461**	$437
4.4-4.9	Other Gasification Equipment	$24,056	$10,168	$14,678	$0	$0	**$48,902**	$4,688	$0	$11,449	**$65,038**	$127
	SUBTOTAL 4	**$324,131**	**$10,168**	**$77,944**	**$0**	**$0**	**$412,242**	**$39,014**	**$24,521**	**$65,332**	**$541,109**	**$1,054**
5A	GAS CLEANUP & PIPING	$89,500	$3,812	$77,878	$0	$0	**$171,190**	$16,546	$26,077	$42,894	**$256,707**	$500
5B	CO2 COMPRESSION	$18,339	$0	$11,242	$0	$0	**$29,581**	$2,849	$0	$6,486	**$38,916**	$76
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	$92,027	$0	$6,583	$0	$0	**$98,609**	$9,348	$9,861	$11,782	**$129,600**	$252
6.2-6.9	Combustion Turbine Other	$0	$806	$892	$0	$0	**$1,699**	$159	$0	$557	**$2,415**	$5
	SUBTOTAL 6	**$92,027**	**$806**	**$7,475**	**$0**	**$0**	**$100,308**	**$9,507**	**$9,861**	**$12,339**	**$132,015**	**$257**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	$33,631	$0	$4,782	$0	$0	**$38,414**	$3,652	$0	$4,207	**$46,272**	$90
7.2-7.9	SCR System, Ductwork and Stack	$3,377	$2,407	$3,153	$0	$0	**$8,938**	$829	$0	$1,589	**$11,355**	$22
	SUBTOTAL 7	**$37,008**	**$2,407**	**$7,935**	**$0**	**$0**	**$47,351**	**$4,481**	**$0**	**$5,796**	**$57,628**	**$112**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$26,321	$0	$4,278	$0	$0	**$30,600**	$2,935	$0	$3,353	**$36,888**	$72
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	$9,952	$903	$6,987	$0	$0	**$17,843**	$1,623	$0	$3,868	**$23,333**	$45
	SUBTOTAL 8	**$36,274**	**$903**	**$11,266**	**$0**	**$0**	**$48,442**	**$4,558**	**$0**	**$7,221**	**$60,222**	**$117**
9	COOLING WATER SYSTEM	$10,387	$9,859	$8,527	$0	$0	**$28,773**	$2,673	$0	$6,406	**$37,852**	$74
10	ASH/SPENT SORBENT HANDLING SYS	$19,651	$1,481	$9,750	$0	$0	**$30,882**	$2,963	$0	$3,691	**$37,536**	$73
11	ACCESSORY ELECTRIC PLANT	$31,778	$12,519	$24,431	$0	$0	**$68,728**	$5,909	$0	$14,164	**$88,801**	$173
12	INSTRUMENTATION & CONTROL	$11,157	$2,052	$7,188	$0	$0	**$20,397**	$1,849	$1,020	$3,877	**$27,142**	$53
13	IMPROVEMENTS TO SITE	$3,416	$2,014	$8,429	$0	$0	**$13,859**	$1,368	$0	$4,568	**$19,796**	$39
14	BUILDINGS & STRUCTURES	$0	$6,693	$7,589	$0	$0	**$14,282**	$1,300	$0	$2,555	**$18,136**	$35
	TOTAL COST	**$722,212**	**$67,672**	**$295,478**	**$0**	**$0**	**$1,085,363**	**$102,090**	**$61,479**	**$197,964**	**$1,446,895**	**$2,817**

Exhibit 3-70 Case 4 Total Plant Cost Details

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$3,737	$0	$1,826	$0	$0	$5,563	$498	$0	$1,212	$7,273	$14
1.2	Coal Stackout & Reclaim	$4,829	$0	$1,171	$0	$0	$5,999	$526	$0	$1,305	$7,830	$15
1.3	Coal Conveyors & Yd Crush	$4,489	$0	$1,158	$0	$0	$5,648	$496	$0	$1,229	$7,372	$14
1.4	Other Coal Handling	$1,175	$0	$268	$0	$0	$1,443	$126	$0	$314	$1,883	$4
1.5	Sorbent Receive & Unload	$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
1.7	Sorbent Conveyors	$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
1.9	Coal & Sorbent Hnd.Foundations	$0	$2,644	$6,612	$0	$0	$9,256	$887	$0	$2,029	$12,172	$24
	SUBTOTAL 1.	**$14,229**	**$2,644**	**$11,035**	**$0**	**$0**	**$27,908**	**$2,533**	**$0**	**$6,088**	**$36,529**	**$71**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying (incl. w/2.3)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.2	Prepared Coal Storage & Feed	$1,596	$382	$250	$0	$0	$2,228	$190	$0	$484	$2,902	$6
2.3	Slurry Prep & Feed	$21,768	$0	$9,681	$0	$0	$31,449	$2,808	$0	$6,851	$41,108	$80
2.4	Misc.Coal Prep & Feed	$877	$639	$1,914	$0	$0	$3,430	$315	$0	$749	$4,495	$9
2.5	Sorbent Prep Equipment	$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
2.7	Sorbent Injection System	$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
2.9	Coal & Sorbent Feed Foundation	$0	$3,411	$2,800	$0	$0	$6,211	$575	$0	$1,357	$8,143	$16
	SUBTOTAL 2.	**$24,241**	**$4,431**	**$14,646**	**$0**	**$0**	**$43,318**	**$3,889**	**$0**	**$9,441**	**$56,648**	**$110**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	Feedwater System	$2,834	$4,868	$2,570	$0	$0	$10,272	$952	$0	$2,245	$13,468	$26
3.2	Water Makeup & Pretreating	$709	$74	$396	$0	$0	$1,179	$112	$0	$387	$1,679	$3
3.3	Other Feedwater Subsystems	$1,551	$524	$472	$0	$0	$2,546	$229	$0	$555	$3,330	$6
3.4	Service Water Systems	$406	$835	$2,899	$0	$0	$4,140	$404	$0	$1,363	$5,907	$12
3.5	Other Boiler Plant Systems	$2,177	$843	$2,090	$0	$0	$5,111	$485	$0	$1,119	$6,714	$13
3.6	FO Supply Sys & Nat Gas	$313	$591	$551	$0	$0	$1,456	$140	$0	$319	$1,915	$4
3.7	Waste Treatment Equipment	$991	$0	$604	$0	$0	$1,595	$155	$0	$525	$2,276	$4
3.8	Misc. Power Plant Equipment	$1,094	$146	$562	$0	$0	$1,802	$174	$0	$593	$2,569	$5
	SUBTOTAL 3.	**$10,074**	**$7,882**	**$10,144**	**$0**	**$0**	**$28,101**	**$2,651**	**$0**	**$7,106**	**$37,858**	**$74**
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries (E-GAS)	$114,050	$0	$63,266	$0	$0	$177,316	$16,295	$24,521	$33,478	$251,609	$490
4.2	Syngas Cooling	w/4.1	$0	w/ 4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.3	ASU/Oxidant Compression	$186,025	$0	w/equip.	$0	$0	$186,025	$18,031	$0	$20,406	$224,461	$437
4.4	LT Heat Recovery & FG Saturation	$24,056	$0	$9,145	$0	$0	$33,201	$3,240	$0	$7,288	$43,730	$85
4.5	Misc. Gasification Equipment	w/4.1&4.2	$0	w/4.1&4.2	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Flare Stack System	$0	$1,643	$669	$0	$0	$2,312	$222	$0	$507	$3,041	$6
4.8	Major Component Rigging	w/4.1&4.2	$0	w/4.1&4.2	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Gasification Foundations	$0	$8,525	$4,864	$0	$0	$13,389	$1,226	$0	$3,654	$18,268	$36
	SUBTOTAL 4.	**$324,131**	**$10,168**	**$77,944**	**$0**	**$0**	**$412,242**	**$39,014**	**$24,521**	**$65,332**	**$541,109**	**$1,054**

Exhibit 3-70 Case 4 Total Plant Cost Details (Continued)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
5A	GAS CLEANUP & PIPING											
5A.1	Double Stage Selexol	$70,224	$0	$59,586	$0	$0	$129,810	$12,554	$25,962	$33,665	$201,991	$393
5A.2	Elemental Sulfur Plant	$10,291	$2,051	$13,278	$0	$0	$25,620	$2,489	$0	$5,622	$33,730	$66
5A.3	Mercury Removal	$1,302	$0	$991	$0	$0	$2,294	$222	$115	$526	$3,156	$6
5A.4	Shift Reactors	$7,138	$0	$2,873	$0	$0	$10,011	$960	$0	$2,194	$13,164	$26
5A.5	Particulate Removal	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
5A.6	Blowback Gas Systems	$545	$306	$172	$0	$0	$1,023	$98	$0	$224	$1,345	$3
5A.7	Fuel Gas Piping	$0	$723	$506	$0	$0	$1,229	$114	$0	$269	$1,612	$3
5A.9	HGCU Foundations	$0	$732	$472	$0	$0	$1,204	$111	$0	$394	$1,709	$3
	SUBTOTAL 5A.	**$89,500**	**$3,812**	**$77,878**	**$0**	**$0**	**$171,190**	**$16,546**	**$26,077**	**$42,894**	**$256,707**	**$500**
5B	CO2 COMPRESSION											
5B.1	CO2 Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5B.2	CO2 Compression & Drying	$18,339	$0	$11,242	$0	$0	$29,581	$2,849	$0	$6,486	$38,916	$76
	SUBTOTAL 5B.	**$18,339**	**$0**	**$11,242**	**$0**	**$0**	**$29,581**	**$2,849**	**$0**	**$6,486**	**$38,916**	**$76**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	$92,027	$0	$6,583	$0	$0	$98,609	$9,348	$9,861	$11,782	$129,600	$252
6.2	Open	$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
6.9	Combustion Turbine Foundations	$0	$806	$892	$0	$0	$1,699	$159	$0	$557	$2,415	$5
	SUBTOTAL 6.	**$92,027**	**$806**	**$7,475**	**$0**	**$0**	**$100,308**	**$9,507**	**$9,861**	**$12,339**	**$132,015**	**$257**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	$33,631	$0	$4,782	$0	$0	$38,414	$3,652	$0	$4,207	$46,272	$90
7.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$0	$1,731	$1,235	$0	$0	$2,966	$260	$0	$645	$3,871	$8
7.4	Stack	$3,377	$0	$1,269	$0	$0	$4,645	$445	$0	$509	$5,599	$11
7.9	HRSG,Duct & Stack Foundations	$0	$677	$650	$0	$0	$1,326	$123	$0	$435	$1,885	$4
	SUBTOTAL 7.	**$37,008**	**$2,407**	**$7,935**	**$0**	**$0**	**$47,351**	**$4,481**	**$0**	**$5,796**	**$57,628**	**$112**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$26,321	$0	$4,278	$0	$0	$30,600	$2,935	$0	$3,353	$36,888	$72
8.2	Turbine Plant Auxiliaries	$182	$0	$417	$0	$0	$599	$59	$0	$66	$724	$1
8.3	Condenser & Auxiliaries	$4,762	$0	$1,521	$0	$0	$6,284	$601	$0	$688	$7,573	$15
8.4	Steam Piping	$5,008	$0	$3,523	$0	$0	$8,531	$733	$0	$2,316	$11,580	$23
8.9	TG Foundations	$0	$903	$1,526	$0	$0	$2,429	$230	$0	$798	$3,457	$7
	SUBTOTAL 8.	**$36,274**	**$903**	**$11,266**	**$0**	**$0**	**$48,442**	**$4,558**	**$0**	**$7,221**	**$60,222**	**$117**

Exhibit 3-70 Case 4 Total Plant Cost Details (Continued)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$7,196	$0	$1,401	$0	$0	$8,597	$819	$0	$1,412	$10,828	$21
9.2	Circulating Water Pumps	$1,877	$0	$136	$0	$0	$2,013	$170	$0	$327	$2,510	$5
9.3	Circ.Water System Auxiliaries	$157	$0	$22	$0	$0	$179	$17	$0	$29	$226	$0
9.4	Circ.Water Piping	$0	$6,545	$1,697	$0	$0	$8,241	$745	$0	$1,797	$10,783	$21
9.5	Make-up Water System	$384	$0	$549	$0	$0	$933	$90	$0	$205	$1,227	$2
9.6	Component Cooling Water Sys	$773	$924	$657	$0	$0	$2,354	$221	$0	$515	$3,090	$6
9.9	Circ.Water System Foundations	$0	$2,391	$4,064	$0	$0	$6,455	$612	$0	$2,120	$9,187	$18
	SUBTOTAL 9.	**$10,387**	**$9,859**	**$8,527**	**$0**	**$0**	**$28,773**	**$2,673**	**$0**	**$6,406**	**$37,852**	**$74**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Slag Dewatering & Cooling	$17,164	$0	$8,464	$0	$0	$25,628	$2,462	$0	$2,809	$30,900	$60
10.2	Gasifier Ash Depressurization	w/10.1	w/10.1	w/10.1	$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
10.4	High Temperature Ash Piping	$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
10.6	Ash Storage Silos	$564	$0	$613	$0	$0	$1,177	$114	$0	$194	$1,485	$3
10.7	Ash Transport & Feed Equipment	$756	$0	$182	$0	$0	$938	$88	$0	$154	$1,180	$2
10.8	Misc. Ash Handling Equipment	$1,168	$1,431	$427	$0	$0	$3,026	$288	$0	$497	$3,811	$7
10.9	Ash/Spent Sorbent Foundation	$0	$50	$63	$0	$0	$112	$11	$0	$37	$160	$0
	SUBTOTAL 10.	**$19,651**	**$1,481**	**$9,750**	**$0**	**$0**	**$30,882**	**$2,963**	**$0**	**$3,691**	**$37,536**	**$73**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$924	$0	$914	$0	$0	$1,839	$176	$0	$201	$2,216	$4
11.2	Station Service Equipment	$4,676	$0	$421	$0	$0	$5,097	$470	$0	$557	$6,124	$12
11.3	Switchgear & Motor Control	$8,644	$0	$1,572	$0	$0	$10,216	$948	$0	$1,675	$12,838	$25
11.4	Conduit & Cable Tray	$0	$4,015	$13,247	$0	$0	$17,262	$1,670	$0	$4,733	$23,665	$46
11.5	Wire & Cable	$0	$7,672	$5,041	$0	$0	$12,713	$924	$0	$3,409	$17,046	$33
11.6	Protective Equipment	$0	$680	$2,474	$0	$0	$3,153	$308	$0	$519	$3,980	$8
11.7	Standby Equipment	$229	$0	$223	$0	$0	$452	$43	$0	$74	$570	$1
11.8	Main Power Transformers	$17,305	$0	$140	$0	$0	$17,445	$1,319	$0	$2,815	$21,579	$42
11.9	Electrical Foundations	$0	$152	$398	$0	$0	$550	$53	$0	$181	$784	$2
	SUBTOTAL 11.	**$31,778**	**$12,519**	**$24,431**	**$0**	**$0**	**$68,728**	**$5,909**	**$0**	**$14,164**	**$88,801**	**$173**
12	INSTRUMENTATION & CONTROL											
12.1	IGCC Control Equipment	w/4.1	$0	w/4.1	$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
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$1,101	$0	$735	$0	$0	$1,837	$174	$92	$315	$2,418	$5
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$253	$0	$162	$0	$0	$415	$39	$21	$95	$571	$1
12.7	Computer & Accessories	$5,875	$0	$188	$0	$0	$6,063	$557	$303	$692	$7,615	$15
12.8	Instrument Wiring & Tubing	$0	$2,052	$4,196	$0	$0	$6,248	$530	$312	$1,773	$8,863	$17
12.9	Other I & C Equipment	$3,927	$0	$1,907	$0	$0	$5,834	$549	$292	$1,001	$7,676	$15
	SUBTOTAL 12.	**$11,157**	**$2,052**	**$7,188**	**$0**	**$0**	**$20,397**	**$1,849**	**$1,020**	**$3,877**	**$27,142**	**$53**

Exhibit 3-70 Case 4 Total Plant Cost Details (Continued)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
13	IMPROVEMENTS TO SITE											
13.1	Site Preparation	$0	$107	$2,291	$0	$0	$2,398	$238	$0	$791	$3,427	$7
13.2	Site Improvements	$0	$1,906	$2,533	$0	$0	$4,440	$438	$0	$1,463	$6,341	$12
13.3	Site Facilities	$3,416	$0	$3,605	$0	$0	$7,021	$692	$0	$2,314	$10,027	$20
	SUBTOTAL 13.	**$3,416**	**$2,014**	**$8,429**	**$0**	**$0**	**$13,859**	**$1,368**	**$0**	**$4,568**	**$19,796**	**$39**
14	BUILDINGS & STRUCTURES											
14.1	Combustion Turbine Area	$0	$265	$150	$0	$0	$414	$36	$0	$90	$541	$1
14.2	Steam Turbine Building	$0	$2,246	$3,200	$0	$0	$5,445	$501	$0	$892	$6,838	$13
14.3	Administration Building	$0	$870	$631	$0	$0	$1,501	$134	$0	$245	$1,880	$4
14.4	Circulation Water Pumphouse	$0	$163	$86	$0	$0	$250	$22	$0	$41	$312	$1
14.5	Water Treatment Buildings	$0	$592	$578	$0	$0	$1,171	$106	$0	$191	$1,468	$3
14.6	Machine Shop	$0	$445	$305	$0	$0	$750	$67	$0	$122	$939	$2
14.7	Warehouse	$0	$719	$464	$0	$0	$1,183	$105	$0	$193	$1,481	$3
14.8	Other Buildings & Structures	$0	$431	$335	$0	$0	$766	$68	$0	$167	$1,001	$2
14.9	Waste Treating Building & Str.	$0	$963	$1,840	$0	$0	$2,802	$261	$0	$613	$3,676	$7
	SUBTOTAL 14.	**$0**	**$6,693**	**$7,589**	**$0**	**$0**	**$14,282**	**$1,300**	**$0**	**$2,555**	**$18,136**	**$35**
	TOTAL COST	**$722,212**	**$67,672**	**$295,478**	**$0**	**$0**	**$1,085,363**	**$102,090**	**$61,479**	**$197,964**	**$1,446,895**	**$2,817**
	Owner's Costs											
	Preproduction Costs											
	6 Months All Labor										$13,491	$26
	1 Month Maintenance Materials										$2,999	$6
	1 Month Non-fuel Consumables										$385	$1
	1 Month Waste Disposal										$295	$1
	25% of 1 Months Fuel Cost at 100% CF										$1,687	$3
	2% of TPC										$28,938	$56
	Total										**$47,793**	**$93**
	Inventory Capital											
	60 day supply of fuel and consumables at 100% CF										$13,995	$27
	0.5% of TPC (spare parts)										$7,234	$14
	Total										**$21,230**	**$41**
	Initial Cost for Catalyst and Chemicals										$7,371	$14
	Land										$900	$2
	Other Owner's Costs										$217,034	$423
	Financing Costs										$39,066	$76
	Total Overnight Costs (TOC)										**$1,780,290**	**$3,466**
	TASC Multiplier								(IOU, high-risk, 35 year)		1.140	
	Total As-Spent Cost (TASC)										**$2,029,531**	**$3,952**

Exhibit 3-71 Case 4 Initial and Annual Operating and Maintenance Costs

INITIAL & ANNUAL O&M EXPENSES		Cost Base (Jun):	2007
Case 4 - ConocoPhillips 500MW IGCC w/ CO2		Heat Rate-net (Btu/kWh):	10,998
		MWe-net:	514
		Capacity Factor (%):	80

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate(base):	34.65	$/hour
Operating Labor Burden:	30.00	% of base
Labor O-H Charge Rate:	25.00	% of labor

Operating Labor Requirements(O.J.)per Shift:	1 unit/mod.	Total Plant
Skilled Operator	2.0	2.0
Operator	10.0	10.0
Foreman	1.0	1.0
Lab Tech's, etc.	3.0	3.0
TOTAL-O.J.'s	16.0	16.0

	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Cost	$6,313,507	$12.292
Maintenance Labor Cost	$15,271,560	$29.734
Administrative & Support Labor	$5,396,267	$10.507
Property Taxes and Insurance	$28,937,909	$56.342
TOTAL FIXED OPERATING COSTS	**$55,919,243**	**$108.875**

VARIABLE OPERATING COSTS

						$/kWh-net
Maintenance Material Cost					**$28,787,121**	**$0.00800**

Consumables	Consumption Initial Fill	Consumption /Day	Unit Cost	Initial Fill Cost		
Water(/1000 gallons)	0	4,116	1.08	**$0**	**$1,300,111**	**$0.00036**
Chemicals						
MU & WT Chem. (lbs)	0	24,523	0.17	$0	$1,239,310	$0.00034
Carbon (Mercury Removal) (lb)	104,394	143	1.05	$109,631	$43,852	$0.00001
COS Catalyst (m3)	0	0	2,397.36	$0	$0	$0.00000
Water Gas Shift Catalyst (ft3)	6,484	4.44	498.83	$3,234,413	$646,883	$0.00018
Selexol Solution (gal)	300,533	98	13.40	$4,026,613	$384,543	$0.00011
SCR Catalyst (m3)	0	0	0.00	$0	$0	$0.00000
Ammonia (19% NH3) (ton)	0	0	0.00	$0	$0	$0.00000
Claus Catalyst (ft3)	w/equip.	2.00	131.27	$0	$76,827	$0.00002
Subtotal Chemicals				**$7,370,657**	**$2,391,415**	**$0.00066**
Other						
Supplemental Fuel (MBtu)	0	0	0.00	$0	$0	$0.00000
Gases, N2 etc. (/100scf)	0	0	0.00	$0	$0	$0.00000
L.P. Steam (/1000 pounds)	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$0**	$0.00000
Waste Disposal						
Spent Mercury Catalyst (lb.)	0	143	0.42	$0	$17,416	$0.00000
Flyash (ton)	0	0	0.00	$0	$0	$0.00000
Slag (ton)	0	593	16.23	$0	$2,809,802	$0.00078
Subtotal Waste Disposal				**$0**	**$2,827,218**	**$0.00079**
By-products & Emissions						
Sulfur (ton)	0	145	0.00	$0	$0	$0.00000
Subtotal By-products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$7,370,657**	**$35,305,866**	**$0.00981**
Fuel (ton)	0	5,811	38.18	**$0**	**$64,786,772**	**$0.01800**