

Impact of CO₂ Capture on Transport Gasifier IGCC Power Plant

Alexander K. Bonsu (contact) 205-670-5898

akbonsu@southernco.com

Joseph D. Eiland, Benjamin F. Gardner, Charles A.

Powell, and Luke H. Rogers

Southern Company Services

Power Systems Development Facility

31800 Highway 25 N, P.O.B. 1069

Wilsonville, AL 35186

George S. Booras

Electrical Power Research Institute

Ronald W. Breault

National Energy Technology Center

U.S. Department of Energy

Nicola Salazar

Kellogg, Brown and Root, Inc.

Abstract

This paper presents the results of two detailed system and economic studies of an Integrated Gasification Combined Cycle (IGCC) plant based on air-blown and oxygen-blown Transport Gasifier technology with carbon dioxide (CO₂) capture. An earlier paper described four cases where gas cleanup with respect to trace metals, sulfur and mercury was the subject of investigation.

Southern Company is developing the Transport Gasifier and related systems for commercial application in the power industry in conjunction with the U.S. Department of Energy (DOE) and Kellogg, Brown and Root, Inc. (KBR). At the engineering-scale Power Systems Development Facility (PSDF) in Wilsonville, AL, several coals have been successfully gasified in both oxygen-blown and air-blown operation of the Transport Gasifier.

To guide future tests and commercialization of the technologies at the PSDF, a series of conceptual commercial plant designs have been completed in partnership with the DOE and the Electric Power Research Institute (EPRI). Two Transport Gasifier combined cycle cases have been developed to investigate the relative costs and benefits of oxygen-blown or air-blown gasification with CO₂ capture using split flow amine system to reduce the solvent regeneration steam consumption. These cases are both based on a 2x1 GE7FA+e combined cycle fueled by syngas from two Transport Gasifiers using Powder River Basin (PRB) sub-bituminous coal.

The performance and cost for each case has been modeled down to the individual equipment level using PSDF test data, chemical and thermal process modeling software, power plant and process plant costing software, vendor quotes, KBR data, and historical Southern Company information.

1 INTRODUCTION

1.1 PSDF

The Power Systems Development Facility (PSDF) was established in 1995 to lead the United States' effort to develop cost-competitive and environmentally acceptable coal-based power plant technologies. This large-scale pilot facility near Wilsonville, Alabama, is focused specifically on identifying ways to reduce capital cost and increase efficiency of advanced coal-based power generation while meeting strict environmental standards.

At the PSDF, the Transport Gasifier and related systems are being developed for commercial applications by Southern Company, Kellogg Brown & Root, Inc. (KBR), the U.S. Department of Energy (DOE) National Energy Technology Laboratory (NETL), and other industrial participants, currently including the Electric Power Research Institute (EPRI), Siemens Westinghouse Power Corporation, Peabody Energy, the Lignite Energy Council, and Burlington Northern Santa Fe Railway. The Transport Gasifier has been used at the PSDF to gasify several coal types ranging from lignite to bituminous in both oxygen-blown and air-blown operation with excellent results.

The Transport Gasifier offers many advantages over commercially available gasifiers that can lead to successful commercialization. These advantages include high carbon conversion with a variety of fuels, high sulfur capture, a small footprint with a high thermal throughput, a simple and robust mechanical design, low water consumption, and the ability to easily gasify high-ash, high-melting-point coals.

1.2 Studies

To guide future tests and commercialization of the technologies at the PSDF, a series of conceptual commercial plant designs have been completed in partnership with DOE NETL and EPRI. Two Transport Gasification combined cycle cases have been developed to investigate the relative costs and benefits of oxygen-blown or air-blown gasification using a split flow system with MDEA solvent to capture CO₂ (Table 1). The split flow amine system was used to reduce the energy consumption required to regenerate the solvent. To ensure that most of the carbon in the coal is captured in the syngas, water gas shift reaction was applied to convert the CO to CO₂. The two cases are described as Case 5 and 6 because this report is part of an economic study of 6 cases. The first four cases

have been described in a paper presented at the Twenty-Second Annual International Pittsburgh Coal Conference held on September 13, 2005 (1).

The list of design basis is as follows:

- Two Transport Gasifiers
- Low-sulfur Powder River Basin (PRB) coal, fed dry
- 2x1 combined cycle plant consisting of 2 GE7FA+e gas turbines and an 1800 psi reheat steam cycle
- Dry ash removal from the syngas at 550°F by metal filter elements
- Selective Catalytic Reduction (SCR) in the Heat Recovery Steam Generator (HRSG) to control nitrogen oxides (NO_x)
- Sulfur in the syngas is captured with the carbon dioxide by the amine systems
- Greenfield site in the southeast United States at 114 feet above sea level with average ambient conditions of 65°F and 60 percent relative humidity

Beyond these similarities, each case is individually optimized for the best cost, performance, and emissions. This leads to differences in results that are sometimes held constant in this type of study, such as net output, emissions, and coal feed rate. Rather than use non-optimum design assumptions to force these outputs to be the same, the resulting differences of each case are allowed to reveal the relative advantages and disadvantages.

Technologies are selected for these configurations that are either commercially available or are available with commercial guarantees in the very near-term. While using developmental technologies such as warm gas cleanup, oxygen membranes, or very high temperature gas filtration would improve the results and possibly change the conclusions, using available technologies and optimizing each case individually results in system designs that are meaningful and pertinent to producing clean power from PRB coal today. Naturally, if other fuels were selected or other end products desired, such as liquid fuels, the systems would be configured differently and the results would be different.

The sulfur in the syngas is captured with the carbon dioxide by the amine system. This is similar to what is being done by Dakota Gasification at the Great Plains Synfuels Plant (2). The carbon dioxide sold to PanCanadian Petroleum Limited for enhanced oil recovery in Saskatchewan, Canada, contains 1 percent H₂S, by volume (3). This approach is only acceptable if the CO₂ is to be sequestered or used for enhanced oil

recovery. If it were to be used for other purposes such as in the beverage or food industry, a separate sulfur removal process would be required.

The performance and cost for each case has been modeled down to the individual equipment level using PSDF test data, chemical and thermal process modeling software, power plant and process plant costing software, vendor quotes, KBR data, and historical Southern Company plant construction and operating information.

The assumptions and calculations in these studies are representative of an Nth plant. For the configurations selected, the technologies are assumed to be mature. For example, no first-of-a-kind costs are added to the capital and no penalty is added to the availability calculations for first-of-a-kind plant startup.

	Gasifier Oxidant	Gas Cleanup
Case 5	air	CO ₂ capture
Case 6	oxygen	CO ₂ capture

Table 1: Two cases using air-blown and oxygen-blown Transport Gasifier

2 SYSTEM DESCRIPTIONS

Since the two cases have common features, they will be described together. The differences will be pointed out in the text. The solids handling, preparation and feed system, and the gasification island and the steam cycle will be described briefly since they have been dealt with in detail elsewhere (1). The system description in this paper will be focused on the separation of carbon dioxide from the syngas. These plants are designed from the beginning with CO₂ capture. The major components of the CO₂ capture system are:

- Water gas shift reaction
- MDEA scrubbing
- CO₂ drying and compression

2.1 Solids Handling, Preparation, and Feed Systems

The design coal is sub-bituminous PRB with an as-received higher heating value of 8,760 Btu/lb, 28 percent moisture, and 0.26 percent sulfur. The coal is delivered to the site by unit trains of bottom-dump, rapid-discharge rail cars. The unloading system will unload each train in 3 to 4 hours and consists of trestles, a below-grade receiving hopper, two belt feeders, and an unloading conveyor. A radial pedestal stacker conveyor is used to form a kidney shaped, 15-day live coal pile. There is a 30-day coal storage area adjacent to the live pile.

Coal is reclaimed from the live coal pile by in-ground, vibrating reclaim bins and directed onto the reclaim conveyor, which transports it to the coal crusher. A

crushed coal conveyor then takes the coal to crushed coal silos which feed the coal drying and milling systems.

The roll mill pulverizers incorporate a flash dryer in which hot gas dries the coal to approximately 18 percent moisture. The drying gas is a mixture of air, water vapor, and nitrogen containing less than 11.3 volume percent of oxygen to meet fire code standards. The oxygen content of the drying gas entering the pulverizer is monitored and nitrogen added as necessary.

A screw conveyor feeds crushed coal from a storage silo to its dedicated pulverizer. The pulverized coal and drying gas passes up the drying column and enters a cyclone, where the majority of the coal is removed and falls into a surge bin. The dusty gas then flows to a baghouse where the coal dust is separated and discharged into the same surge bin. An induced-draft fan before the bag house and a blower before the pulverizer drive the gas through the drying circuit.

A series of water-cooled shell-and-tube heat exchangers supplied with cooling water are used to cool the drying gas below the dew point to condense and knock out the moisture picked up in the pulverizer. The condensed water withdrawn from the knock-out drum is sent to the water treatment plant, since it includes coal dust transmitted through the bag house. The cooled gas is reheated in shell-and-tube heaters using medium pressure steam and the hot gas is recirculated back to the pulverizer to dry more coal. Steam heating is preferred as it avoids the operating cost associated with fuel-fired burners and it also minimizes the amount of moisture present in the drying gas and improves drying efficiency.

The coal feed system to the gasifier consists of a surge bin that receives the prepared coal, a lock vessel, a feed vessel, and a rotary feeder with a vertical axis. The coal is transported into the gasifier by air in Cases 5 and by recycle syngas in Cases 6 via dilute-phase conveyors. The Powder River Basin coal composition (by weight percent) and heat content as fed to the gasifier are shown in Table 2.

If there is not a bed of ash in the gasifier when it first starts up, then a bed of solids is first fed to the gasifier by the startup solids feed system. Sand is purchased already dried and prepared to a top size of 500 microns for this purpose, and recovered gasifier bed material can also be used. A single feed system introduces the sand into each gasifier, consisting of a surge bin, a lock vessel, a feed vessel, and two rotary feeders. The startup solids are transported into the gasifiers using air in dilute-phase conveyors.

	<u>Weight %</u>
C	58.4
H	3.8
O	13.0
N	0.8
S	0.3
Ash	5.8
H ₂ O	17.9
Cl	0.012
F	0.004
Hg	8*10 ⁻⁶
	<u>Btu/lb</u>
HHV	9,852
LHV	9,305

Table 2: Coal properties, as fed to the gasifier

2.2 Gasification

The Transport Gasifier is the heart of the gasification island. It operates as a pressurized, circulating fast-fluidized bed gasifier and consists of simple refractory-lined pipe sections. The design operating temperature and pressure are approximately 1,800°F and 435 psia, respectively. Thermal expansion is accommodated without recourse to expansion joints.

The fuel used is low-sulfur PRB sub-bituminous coal. Air and steam are used to gasify the coal to achieve 97% carbon conversion. In Case 6, the oxygen is produced using a conventional cryogenic Air Separation Unit (ASU). This is a high pressure ASU and purity is 95% oxygen. There is no limestone feed to the gasifier to capture sulfur in either case. The sulfur in the form of H₂S is removed with the CO₂. The COS present in the syngas leaving the gasifier is subsequently hydrolyzed to H₂S and CO₂ in the water gas shift reactors.

Most of the entrained gasification ash is removed from the syngas in the disengager and cyclone and recirculated back to the gasifier. The syngas and remaining particulates are cooled in the primary syngas cooler by raising high pressure saturated steam, and then further cooled in the secondary syngas cooler by raising medium pressure saturated steam. The entrained gasification ash is then captured when the syngas passes through banks of metal filter elements in a vessel called a particulate control device (PCD). The temperature of the PCD is maintained at 550°F. This is the desired temperature for the water gas shift reaction which follows the PCD.

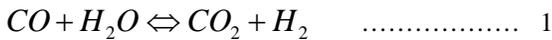
The composition of the syngas leaving the gasifier in Cases 5 and 6 is given in Table 3.

	Case 5	Case 6
	<u>Vol. %</u>	
CH ₄	2.14	2.57
CO	22.25	31.48
CO ₂	7.92	18.69
H ₂	11.30	28.49
HCN	0.02	0.04
H ₂ O	5.44	17.24
H ₂ S	0.07	0.12
N ₂	50.70	1.10
NH ₃	0.16	0.27
Molecular Weight	25.6	21.54
Flow Rate, lb/hr	1,999,000	1,061,000
	<u>Btu/SCF</u>	
LHV	119.1	198.1
HHV	129.4	222.9

Table 3: Syngas composition, flow rate and heating value exiting the gasifier

2.3 Water Gas Shift Reaction System

After the PCD, the syngas is fed to fixed-bed catalytic reactors to convert most of the CO to CO₂. When sub-bituminous coal is gasified, nearly one-half to three-quarters of the carbon is converted to CO, depending on whether the gasification process is oxygen-blown or air-blown, respectively (Table 3). Since the CO is a major component of the energy in the syngas, it must be converted to CO₂ and hydrogen, by reacting it with steam in order to ensure that most of the carbon containing compounds can be removed prior to burning the syngas to produce power. The reaction of CO with steam to produce hydrogen is often referred to as water gas shift reaction (WGSR):



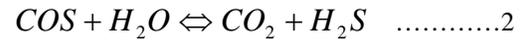
This reaction is exothermic and reversible. This results in a loss of heating value so the coal feed system, gasifiers, ASU, etc. need to be oversized for the CO₂ capture cases relative to the non-capture cases.

The equilibrium conversion is temperature dependent and it is favored by lower temperatures. However, at low temperatures where the equilibrium conversion is high, the reaction rate is low. The reaction is thus usually carried out in two steps. The first part of the reaction is maintained at a high temperature (650-750 °F) to convert the bulk of the CO to CO₂ at a relatively fast reaction rate. The second reactor is maintained at a relatively low temperature (300-400 °F) to complete the rest of the reaction. The catalysts that are used in the two step

approach, iron-chrome catalyst for the high temperature reaction and a mixture of copper oxide and zinc oxide for the low temperature reaction, are susceptible to sulfur poisoning by reduced sulfur compounds such as H₂S and COS. Since the syngas in Cases 5 and 6 contains these sulfur compounds, the catalysts for the two-step approach can not be used. The catalyst for the two step approach is referred to as “sweet shift” catalyst. Instead of the sweet shift, a cobalt-molybdenum catalyst, that operates at the medium temperature of 550 °F and that is not susceptible to reduced sulfur compounds poisoning, is used. In fact this catalyst requires about 100 ppm H₂S in the syngas for it to be effective and it is referred to as “sour shift” catalyst.

Like all WGSR catalysts, the cobalt-molybdenum catalyst is poisoned by particulates, chlorides and condensed water. In both Cases 5 and 6 the syngas filter protects the catalyst from particulates, an upstream guard bed protects it from chlorides, and the temperature is maintained high enough to prevent water condensation.

The COS in the syngas is hydrolyzed to H₂S over the WGSR catalyst according to the following equation:



This enables practically all of the sulfur to be captured in the downstream amine process. The HCN in the syngas is hydrolyzed to NH₃ over the WGSR catalyst as follows:



The water vapor in the syngas is not enough to achieve a high conversion of CO, so extra steam is added in both cases. Since WGSR is an equilibrium reaction (Equ 2), the conversion of CO to CO₂ for a fixed H₂ content is a function of the steam concentration as determined by the equilibrium constant, K_p:

$$K_p = \frac{CO_2 * H_2}{CO * H_2O} \quad \dots\dots\dots 4$$

The equilibrium constant is a function of temperature, so at a given temperature, the higher the conversion of CO desired, the higher the amount of steam that must be added. Initially, when the WGSR was modeled using a single reactor the amount of steam required was very high. By modeling the reaction in three reactors in series and optimizing the conversion in each reactor, the steam consumption was reduced by about two thirds from the single reactor steam flows.

The water gas shift reaction is exothermic and the catalyst temperature limit is 900 °F, so heat exchangers are required before the second and third reactors in series to cool the syngas to the desired inlet temperature of 550 °F.

2.4 Amine System

After the WGSR the syngas is cooled to near ambient temperature by steam cycle condensate and cooling water for Case 5. For Case 6, the syngas from the WGSR is cooled by raising low pressure steam before being further cooled to near ambient temperature with condensate and cooling water. The cooled syngas in both cases is then passed through a knock-out drum to separate the condensed water. Next, mercury is removed from the syngas in a bed of sulfur-impregnated activated carbon.

The syngas is then introduced to an MDEA scrubbing system to separate CO₂ and other acid gases such as H₂S. To reduce the solvent regeneration energy cost and to ensure high CO₂ removal efficiency, a split stream scrubbing system shown is used, as shown in Figure 3. The raw syngas enters the bottom of the absorber where the bulk of the CO₂ is scrubbed with nearly 80% of the semi-lean solution produced by stripping the rich solution leaving the bottom of the absorber with the acid gas leaving the lean regenerator. The rest of the semi-lean solution is fed to the lean regenerator where all the CO₂ is stripped to produce lean amine solution that is fed to the top of the absorber as a polishing agent to complete the scrubbing of the CO₂ from the syngas. In the lean regenerator, the CO₂ and other acid gases such as H₂S are stripped from the solvent with steam generated by the reboiler at the bottom. The reboiler is the major energy user and it is operated at 240-250 °F. The vapor and the acid gases leaving the top of the lean regenerator are fed to the bottom of the semi-lean regenerator to provide a source of heat for stripping the bulk of the CO₂ from the rich amine exiting the bottom of the absorber. The final vapor and acid gases leaving the semi-lean regenerator are cooled by condenser before being fed to a separator to knockout the moisture, and then recycled to the semi-lean regenerator. The acid gas leaving the top of the separator is mainly CO₂ that is ready for recompression and storage or transportation. The absorber is operated at the high pressure of the syngas (404 psia) to favor the CO₂ absorption while the semi-lean regenerator and the lean regenerator are operated at a much reduced pressure (25 psia) to enhance the CO₂ stripping.

To maintain the integrity of the scrubbing solution, a slip stream of lean solvent equivalent to about 10% of the total solvent flow is filtered through particulate and carbon filters. The particulate filter prevents solid build-

up while the carbon filter removes formates and carbamates generated by the small amount of CO remaining in the syngas. The two-stage filtration is essential in minimizing foaming and corrosion. The particulate filter elements are made from virgin cotton or inert polymer fiber. Treated fibers tend to lose their coating into the MDEA system, causing foaming (4). The filter elements are 3" diameter by 36" long and are sized to handle 4 gpm. For the carbon filter, the residence time is 15 minutes and the linear velocity is 5 gpm/ft² (5).

The type of amine compound that is chosen for this process is a modified form of methyl diethanolamine (MDEA = CH₃N(C₂H₄OH)₂) CS-2020 made by GAS/SPEC. It is formulated to provide deep removal of CO₂. Although unconfirmed, it is speculated to be 35% MDEA plus 15% DEA with the balance water. In general, MDEA has the highest chemical and thermal stability compare to other amines. It does not form stable salts or degradable products with any of the major compounds in the syngas with the exception of CO whose concentration in the syngas must be kept below 1 v% (5). Since the syngas is virtually free of particulates and acid gases such as SO₂ and NO₂ capable of forming heat stable salts, it is expected that no reclaiming of the MDEA solvent will be necessary. However, if it is required then technologies such as ion exchange resin, electrochemical cells and vacuum distillation can be applied.

Although MDEA is less reactive than mono ethanolamine (MEA) with acid gases at low pressures, it is quite reactive at high pressures. In addition, since it has lower heat of reaction compared to MEA, it requires less energy to regenerate it at low pressures. Furthermore, MDEA can be used in concentrations as high as 60% compared to 20% for MEA (6). Also, the absorption capacity of MDEA can be as high 0.7 mole acid gas/mole amine compared to 0.35 mole acid gas/mole amine for MEA (7). Hence, MDEA requires less circulation rate to remove the same amount of acid gas compared to MEA, further reducing the energy required to regenerate the solvent.

Sour water condenses from the syngas as it is cooled below the dew point before entering the MDEA split-stream scrubber. The water dissolves almost all the nitrogenous compounds, chloride, and fluoride present as well as lesser amounts of H₂S, COS, CO, and CO₂. The sour water is removed in the low temperature syngas coolers and in a knock-out drum and passed to the sour water treatment plant along with other minor sour water streams. Off gases from this conventional stripping process are compressed and recycled to the process. The ammonia is recovered as anhydrous ammonia for use in the SCR (described below) and for byproduct sales. The recycled off gases, along with additional recycle syngas taken if needed from the main syngas stream before sulfur

treatment, are used for gasifier aeration; filter element back pulsing; and, in the oxygen-blown cases, coal conveying.

The carbon dioxide-rich stream stripped from the amine at 25 psia in the regenerators is dried and compressed to pipeline pressure by the CO₂ compression system. The gas entering the system contains 3.3% water by weight and trace amounts of other syngas constituents.

Any entrained liquids in the gas are first removed in an inlet suction drum. The gas is then compressed through three stages of compression, intercooling, and liquid knockout to 515 psia. It is then dried to a -40°F dewpoint (100 ppmv) in a TEG dehydration package. The gas then enters a fourth stage of compression where it is compressed to 1,210 psia. At this pressure it is above its critical conditions. The gas is then cooled in an aftercooler, but no knockout is required since it is in the dense phase. It then enters the multi-stage CO₂ pump and is discharged at 2,200 psia.

Steam and cooling water are supplied to the CO₂ compression system from the medium-pressure steam header and the cooling tower, respectively.

2.5 Combined Cycle

The combined cycle is built around the GE 7FA+e gas turbine, modified for operation on syngas. The modifications include replacing the standard dry low-NO_x combustor cans with flame diffusion combustors to prevent flashback. The diffusion combustors are expected to be suitable for burning hydrogen fuel.

The gas turbine is fueled by natural gas when syngas is not available, both during gasifier outages and gasifier start-up. When the gas turbine is firing natural gas, water is injected into the combustion cans to limit thermal NO_x formation. This is not necessary when firing syngas since the dilute syngas and the diluent nitrogen (in Case 6) keep the flame temperature low.

An evaporative cooling system at the gas turbine compressor inlet is used at high ambient temperatures to cool the turbine inlet air. All other standard combined cycle auxiliary equipment such as generator cooling, lube oil pumps and water wash systems is included.

The HRSG is a dual-pressure unit with single reheat having a main steam pressure of 1800 psia. Because of lower than usual CT exhaust temperatures, the main and reheat steam temperatures are 980°F/970°F for Case 5 and 960°F/960°F for Case 6. Condensate from the steam cycle is warmed in the gasification island before it enters the cold end of the HRSG. The water is heated at low pressure and then sent to the deaerator, which is

integrated with the low-pressure steam drum. Some of the water from the deaerator is pumped to medium pressure and then sent to the medium pressure steam drum in the gasification island, while the remainder of the water is pumped to high pressure and sent to the high pressure economizers in the HRSG. Most of the economized water is sent to the gasification island high pressure steam drum with the rest sent to the HRSG high pressure steam drum. Desuperheating water is taken from the high pressure and medium pressure feedwater pump outlets when needed.

Saturated steam from the gasification island high pressure steam drum is returned to the HRSG high pressure steam drum so that any entrained water can be removed by the steam drum demisters. The combined steam flow is superheated in the HRSG and then sent to the high pressure section of the steam turbine. Blowdown from the high pressure steam drums is flashed at low pressure, and the steam is sent to the deaerators and the remaining water to treatment. A portion of the cold reheat steam exiting the high pressure section is sent to the gasification island for injection in the gasifier, for use in the water gas shift process, and for drying both coal and carbon dioxide. The remaining cold reheat steam is reheated in the HRSG and then expanded in the medium pressure section of the steam turbine.

Low pressure steam is taken from the crossover as the steam is sent to the low pressure steam turbine and then to the ASU in Case 6 and to the Sour Water Treatment system in both cases. Steam is also extracted from the low pressure turbine and used to regenerate the amine in the amine system. These extractions consume about half of the steam in Case 5 and most of the steam in Case 6 before it exhausts from the low pressure turbine at 1 psia and is condensed by water from a wet mechanical draft cooling tower. The condensate is then sent to the gasification island where it is used for process cooling and then combined with condensate from the other steam uses. The cooling tower also supplies cooling water for other process areas as needed, such as process air compressor intercoolers, ASU, amine system, CO₂ compressor, and coal mill systems.

The HRSG exhaust temperature is maintained at least 20°F above the worst-case acid dewpoint temperature of the flue gas in each case by selection of the HRSG steam drum pinches. A selective catalytic reduction system is installed in the HRSG to reduce NO_x emissions. The anhydrous ammonia reagent used in the SCR is produced from ammonia recovered from the sour-water treatment plant.

When the gasification island is not operating and generating steam, the HRSG alone must raise all of the high pressure steam. In this mode of operation, a duct

burner upstream of the HRSG evaporator section fires natural gas to boost steam flow and pressure.

3 METHODOLOGY

The performance calculations begin with individual gasifier heat and material balances for each case, based on test results of the Transport Gasifier at the PSDF. The gasification-related components and the WGSR system are modeled using Aspen Plus® from AspenTech, Inc. The MDEA scrubbing system was modeled with ProMax process simulation software developed by Bryan Research & Engineering, Inc. for amine systems. KBR provided the composition of the sour water condensed from the syngas as it is cooled and data for the sour water treatment systems. The combined cycle in each case is modeled using Thermoflex from Thermoflow, Inc, with the gas turbine model calibrated with performance calculations from GE Power Systems.

Obtaining performance calculations and data from outside sources requires a significant lead time, and in many cases the system information changed after calculations were requested. When necessary, data from outside sources was adjusted or scaled to match the new process conditions or the data was used to calibrate commercial models which then generated revised calculations.

Gasification-related, solids handling, WGSR, MDEA scrubbing and BOP equipment are sized using Southern Company-proprietary routines, by using Kbase™ software from AspenTech, or by equipment vendors. Combined cycle equipment is sized using Thermoflex or GT PRO software, both from Thermoflow, Inc. Equipment costs are determined from vendor quotes, Southern Company-proprietary databases, Kbase, or Thermoflow's PEACE.

The capital costs include equipment, labor, materials, indirect construction costs, engineering, contingencies, and land. Land is valued at \$5,100 per acre. Sales tax is 5 percent and freight is 2 percent of the equipment cost. Engineering is 7 percent of the total field cost. An overall contingency factor of 10 percent is applied to the estimate. Allowance for Funds Used During Construction (AFUDC) is calculated from monthly projected expenditures using current market rates over a cash flow period of sixty months. The result is approximately 14 percent of the overnight project cost. The capital costs are reported in January 2003 dollars.

The costs are representative of an Nth plant, so no first-of-a-kind costs are included. Appropriate equipment sparing is included in the capital costs, such as 4 x 50% feed water pumps and 7 x 17% coal mill systems.

The Operations and Maintenance (O&M) cost estimates for each case are developed from the component level rather than by applying rule-of-thumb factors to the overall capital cost of the plant. For each case, detailed O&M models are developed that incorporate a full listing of plant equipment along with algorithms to estimate the operating manpower and requirements for maintenance (with scheduled capital costs) and expendables. These models draw on PSDF operating experience, Southern Company's operating experience with coal- and natural gas-fired power plants, and KBR's experience from the operation of their fluid catalytic cracking units.

4 RESULTS AND DISCUSSION

4.1 Water Gas Shift Reaction

The syngas from the gasifier (Table 2) contains about 2% methane which is not affected by the WGSR and the MDEA scrubbing, so it ends up in the gas turbine where it is combusted to CO₂. Therefore to achieve 90% removal of the carbon in the syngas, 95% of the CO must be converted to CO₂ in Case 5 and 96% in Case 6. These high conversions are also required to keep the CO concentration in the syngas below 1% in order to avoid degrading the amine solution in the MDEA scrubber.

Based on the syngas volumetric flow rate and standard sizing parameters for the reactors, 6 parallel WGSR trains are required in Case 5 and 3 parallel trains in Case 6. Therefore the total number of WGS reactors is 18 in Case 5 and 9 in Case 6.

In both cases, the bulk of the CO conversion occurs in the first reactor where the CO concentration is lowered from 22.25% to 3.45% in Case 5 and from 31.53% to 4.11% in Case 6. The last two reactors in series are polishing stages required to bring the CO concentration below 1%. The syngas compositions leaving the WGS reactors and the details of the WGS reactor performance are given in Table 4, 5A and 5B, respectively.

	Case 5	Case 6
	<u>Vol. %</u>	
CH ₄	1.60	1.56
CO	0.80	0.92
CO ₂	21.75	29.60
H ₂	24.07	35.42
HCN	0.01	0.03
H ₂ O	14.00	31.57
H ₂ S	0.05	0.08
N ₂	37.60	0.66
NH ₃	0.12	0.16
Mole. Wt.	23.69	20.19
	<u>Btu/SCF</u>	

LHV	81.3	111.9
HHV	101.3	145.9

Table 4: Syngas compositions exiting WGSR and entering the MDEA scrubber

	Case 5		
	1	2	3
Reactors #	1	2	3
Trains in parallel	6	6	6
Inlet temperature, °F	550	550	550
Outlet temperature, °F	796	593	556
Feed dry gas, lbmole/hr	73929	87753	90074
Syngas H ₂ O, lbmole/hr	4240	14087	13596
Extra steam, lbmole/hr	23672	1827	1492
Inlet steam/dry gas ratio	0.378	0.181	0.168
Outlet H ₂ O/dry gas ratio	0.160	0.151	0.163
Catalyst volume, ft ³	10200	15000	15000
Dry gas inlet SV, scfh/ft ³	2991	2422	2487
Outlet equilibrium CO, %	3.46	1.16	0.80
Pressure drop, psi	6.28	4.14	4.30
Reactor inlet pres., psia	408	401	396
Conv. in single react., %	80	65	29
Total Conversion, %	95		

Table 5A: WGSR details for Case 5

	Case 6		
	1	2	3
Reactors #	1	2	3
Trains in parallel	6	6	6
Inlet temperature, °F	550	550	550
Outlet temperature, °F	828	597	556
Feed dry gas, lbmole/hr	40666	52973	55036
Syngas H ₂ O, lbmole/hr	8461	23275	23346
Extra steam, lbmole/hr	27121	2135	2517
Inlet steam/dry gas ratio	0.875	0.480	0.470
Outlet H ₂ O/dry gas ratio	0.439	0.424	0.461
Catalyst volume, ft ³	5100	7500	7500
Dry gas inlet SV, scfh/ft ³	2789	2533	2651
Outlet equilibrium CO, %	4.11	1.37	0.92
Pressure drop, psi	8.81	5.80	6.16
Reactor inlet pres., psia	408	398	391
Conv. in single react., %	80	70	33
Total Conversion, %	96		

Table 5B: WGSR details for Case 6

4.2 Amine System

The MDEA split-stream CO₂ capture modeling results are summarized in Table 6. There are three parallel trains for each case. The flow rates and the power consumption listed in the table are for the three trains combined. The amine solution flow rates in Cases 5 and 6 are approximately the same because the amount of CO₂ captured is about the same in both cases. Since the absorber pressure is the same in both cases, the auxiliary

power consumptions for the lean pump and the semi-lean pump are also the same.

When the conventional single stage absorber-stripper was used to model the CO₂ capture from the syngas, the reboiler energy consumption required to regenerate the solvent was very high (1641 Btu/lb CO₂ in Case 5 and 1653 Btu/lb CO₂ in Case 6). By applying the split-stream process, the reboiler energy was reduced by 54% to 751 Btu/lb CO₂ in Case 5 and by 53% to 778 Btu/lb CO₂ in Case 6. However this savings was partially off-set by an increase in pumping power.

Aside from chemical and thermal losses, the MDEA solvent will also be lost mechanically through entrainment and evaporation in absorber. The typical losses of MDEA are about 0.5 lb/MMSCF of product syngas leaving the absorber. In Case 5 and 6 the expected losses will be 2.9 and 1.7 lb/hr, respectively.

The equipment sizes are summarized in Table 7. An interesting finding of these studies was that the equipment sizes for the MDEA scrubber to capture CO₂ were identical for both cases even though the mass flow rate of the syngas leaving the gasifier for Case 5 is almost twice that of Case 6 (see Table 3). The initial calculations were done with PROMAX developed by BRE and these were later confirmed by Ineos, Koch-Glitsch and Sulzer.

Absorption towers are sized in two ways, based on the gas flow and based on the solvent flow rate, and the larger size is selected. In both cases the size was set by the solvent flow rate which, as discussed above, is nearly identical for both. The stripper columns are always sized based on the solvent flow rate, so they're also the same for both cases. There are three trains for each case. In each train there are two absorbers and two semi-lean strippers, making a total of six absorbers and six semi-lean strippers. There is only one lean stripper in each train, making a total of three lean strippers.

The syngas is fed to the gas turbine after the MDEA process for scrubbing the CO₂. The compositions of the syngas leaving the MDEA scrubber and entering the gas turbine in the two cases are given in Table 8. When most of the CO₂ is captured, the H₂S is removed down to ppb levels.

	Case 5	Case 6
Number of trains	3	3
Absorber inlet gas temp., °F	100	100
Absorber outlet gas temp., °F	120	120
Absorber lean temp., °F	120	120
Absorber semi-lean temp., °F	145	145
Absorber rich temp., °F	160	159
Absorber press., psia	404	404

Syngas flow rate, MMSCFD	812	465
CO ₂ removed, lb/hr	934950	902715
Lean MDEA conc., %	50	50
MDEA losses, lb/hr	2.9	1.7
Lean stripper press., psia	28.5	28.5
Lean stripper temp., °F	256	256
Semi-lean stripper pres., psia	26.2	26.2
Semi-lean stripper temp., °F	162	162
Lean solution flow rate, gpm	13800	13818
Semi-lean solution flow, gpm	50955	50955
Lean pump power, hp	4492	4492
Semi-lean pump power, hp	15976	15976
Reboiler heat load, Btu/lb CO ₂	751	778
Rich load*	0.539	0.530
Lean load*	0.010	0.009
Semi-lean load*	0.487	0.484
CO ₂ capture efficiency, %	98.5	98.8
Overall carbon capture eff., %	89.8	91.9

Table 6: MDEA CO₂ capture modeling conditions

* mol acid gas/mol amine

	Ves. #	Case 5	Case 6
Number of trains		3	3
Absorber top diam., ft	6	11	11
Absorb bottom diam., ft	6	22	22
Absorb top ht., ft		47	47
Absorb bottom ht., ft		23	23
Semi-lean strip diam., ft	6	20	20
Semi-lean strip ht., ft		28	28
Lean strip diam., ft	3	21	21
Lean strip ht., ft		52	52

Table 7: MDEA CO₂ capture equipment sizes

	Case 5	Case 6
		<u>Vol. %</u>
CH ₄	2.39	3.70
CO	1.24	2.37
CO ₂	0.03	0.01
H ₂	37.56	91.80
HCN	0.00	0.00
H ₂ O	0.20	0.40
H ₂ S	0.00	0.00
N ₂	58.58	1.72
NH ₃	0.00	0.00
Mol. Wt.	18.03	3.669
		<u>Btu/SCF</u>
LHV	124.7	283.8

HHV	145.2	331.9
-----	-------	-------

Table 8: Syngas compositions exiting MDEA scrubber and entering the combustion turbine

4.3 Performance

The performance of the two configurations is given in Table 9. The air-blown system (Case 5) produces more power than the oxygen-blown (Case 6) primarily because the larger syngas flow raises more high pressure steam in the syngas cooler. The power consumed by the process air compressor in Case 5 (54.3 MW) is less than that consumed by the ASU in Case 6 (63.6 MW), and the energy is used in the air-blown case to send additional gas through the gasifier, WGSR, syngas coolers and MDEA scrubber rather than to separate air into nitrogen and oxygen. The coal feed rate is higher in Case 5 because of the energy required to heat the large amount of gas in the gasifier. The syngas flow is determined by the amount necessary to fire the two gas turbines to their full output.

	Case 5	Case 6	% Diff.
Net power, MW	419.8	344.5	-17.9
Net eff. (HHV), %	28.4	24.9	-12.3
Net eff. (LHV), %	30.1	26.3	-12.6
HR (HHV), Btu/kWh	12,000	13,700	14.2
HR (LHV), Btu/kWh	11,300	13,000	15.0
Coal feed, lb/hr	577,000	541,000	-6.2
GT power, MW	394.0	394.0	0.0

Table 9: Power, efficiency, and heat rate

4.4 Capital Costs

The capital costs are reported in Table 10 in January, 2003, thousand U.S. dollars. Recent escalations in commodity prices such as steel and concrete since January, 2003 are not reflected in the capital costs. The indirect costs, which include expenses such as AFUDC, engineering, project management, construction management, temporary facilities and services during construction, startup, insurance, taxes, and land, are about one-third of the total cost. The total cost and levelized cost for Case 5 are lower than that of Case 6, mainly due to the ASU cost (\$87 million) for Case 6 which was about 9 times that of the air compressor cost (\$10 million) for Case 5. The cost of the water gas shift reactors was about 50% higher for Case 5 (\$30 million) compared to Case 6 (\$20 million) mainly due to the difference in the gas flow rates. However, the cost of the amine system to separate CO₂ from the syngas (\$47 million) and the cost of the compressor for the CO₂ (\$25 million) were the same for both cases since the total amount of CO₂ in both cases were nearly the same.

Area Description	Case 5	Case 6
Indirects	290,000	300,000
Equipment Cost	630,300	681,800
TOTAL	920,300	981,800
TOTAL, \$/kW	2,192	2,850

Table 10: Capital costs in thousand U.S. dollars, January, 2003

4.5 Operating and Maintenance Costs

The O&M costs, reported in Table 11, are distributed between Fixed and Variable O&M according to the EPRI TAG[®] basis (see Reference 4). This basis shifts many costs that would normally be considered variable O&M to fixed O&M, and increases the total O&M by approximately \$1/MWhr for each case because several items are not capitalized in the TAG approach. In the table below, fixed O&M includes labor, training, building and grounds maintenance, gas turbine long term service agreements, repair costs, spare parts inventory charges, parts for preventive maintenance, outage expenses, operational upgrades, and startup expenses.

The levelized fixed O&M cost for Case 6 is higher than the value for Case 5. The reason is that the total annual fixed cost is higher and the net power output is lower for Case 6 compare to Case 5. The higher total annual cost for Case 6 is due to the extra compressors for oxygen and nitrogen plus the fact that the maintenance of the oxygen plant's compressors and piping is more time consuming and costly due to the extra safety considerations in handling oxygen. To avoid fire, the piping and equipment contacting with the oxygen must be clean. Furthermore, oxygen systems require specialized lubricants, sealers and packing materials. The net power output for Case 6 is lower than that for Case 5 due to the auxiliary power required by the air separation unit and also due to the fact that the steam turbine power output for Case 6 is about 31% lower than that for Case 5.

Variable O&M includes the costs of expendables and natural gas for startup and flare. The total annualized variable O&M cost is higher for Case 5 compared to Case 6. This is mainly due to the higher WGSR steam consumption in Case 5 compare to Case 6 and the higher labor cost for maintaining the larger number of WGSR reactors for Case 5 compare to the smaller number of reactors for Case 6. However, the annual average generation for Case 5 is also higher than Case 6. Hence, the levelized variable cost obtained by dividing the total annual variable O&M cost by the annual average power generation is lower for Case 5 than Case 6.

	Case 5	Case 6
Net power, MW	419.8	344.5

Ann. avg. gen., 10 ⁶ MW-hr	2.747	2.245
Total fixed (annual) costs, MM\$/yr	24.721	25.649
Total variable (annual) Costs, MM\$/yr	3.188	2.927
Fixed O&M, \$/kW-yr	58.969	74.875
Variable O&M, \$/MW-hr	1.161	1.303

Table 11: Operating and maintenance costs

4.6 Cost of Electricity

The cost of electricity is calculated for each case using the following financial assumptions:

- Plant book life – 20 years
- Carrying charge factor – 0.142
- Capacity factor – 80 percent
- Coal cost – \$1.25/MMBtu

The resulting levelized costs for capital, O&M, and fuel, and the cost of electricity for each case are reported in Table 12, in January, 2003 dollars. The values in Table 13 do not all sum to the given totals due to rounding errors. The calculated values for AFUDC that are included in Table 12 are subtracted out of the total costs so that the standard carrying charge factor can be used in the COE calculations for Table 13.

Care 5 shows lower cost of electricity than Case 6 because of advantages in all three COE components: capital, O&M, and fuel.

	Case 5	Case 6
Capital, mills/kWh	39.1	50.8
O&M, mills/kWh	9.9	11.9
Fuel, mills/kWh	15.0	17.1
Total COE, mills/kWh	64.0	79.8

Table 12: Levl. capital, O&M, fuel, and electric. costs

4.7 Key Variables

Major results of Cases 5 and 6 are compared in Table 13 to show the difference between CO₂ capture configurations using air-blown and oxygen-blown gasification. The differences in net output, efficiency, steam turbine output, and auxiliaries are similar to previous comparisons of air-blown and oxygen-blown gasification without CO₂ capture (1). For most of these key variables, Case 5 is more favorable than Case 6. The costs of CO₂ capture and avoidance are lower for the air-blown Case 5 than for the oxygen-blown Case 6. The lower net outputs of the oxygen-blown cases, compared with the air-blown cases, give them higher costs of electricity to start with (see Table 10). Since the CO₂ capture performance and costs are fairly independent of the gasifier oxidant, the air-blown COE advantage carries over into a cost of carbon capture advantage.

With a lower carbon capture cost, lower cost of electricity, higher net output, and higher efficiency, air-blown gasification is clearly superior to oxygen-blown for these arrangements of Transport Gasification of PRB coal. There is, however, potential for these results to change as improvements are made to both the air-blown and oxygen-blown system configurations.

	Case 5	Case 6	% Diff.
Gasifier oxidant	air	oxygen	
CO ₂ capture	yes	yes	
Net output, MW	419.8	344.5	-17.9
Net eff. (HHV), %	28.4	24.9	-12.3
ST output, MW	190.3	135.4	-28.8
Auxiliaries, MW	166.2	174.5	5.0
Coal feed rate, lb/hr	577,000	541,000	-6.2
Capital cost, million\$	920.3	981.8	6.7
Capital cost, \$/kW	2,192	2,850	30.0
O&M, mills/kWh	9.6	11.6	20.8
COE, mills/kWh	65	81.7	25.7
Carbon capture, %	89.8	91.9	2.3
CO ₂ capt. cost, \$/ton	22	27.9	26.8
CO ₂ avoid. cost, \$/ton	34.3	48.3	40.8

Table 13: Key variables

4.8 Comparing Cases With and Without Carbon Dioxide Capture

Important results for air-blown cases with and without CO₂ capture are shown in Table 14. Case 3, rather than Case 1, is chosen for comparison to Case 5 since it uses syngas cleanup rather than stack gas. In Case 1 the contaminants in the coal such as mercury, sulfur and nitrogen were removed from the flue gas after burning the syngas in the combustion turbine. For Case 3 mercury, sulfur and ammonia were removed from the syngas prior to the combustion turbine, similar to Case 5. This makes Case 3 a good reference plant for Case 5. The details of Case 1 to 4 can be found in the previous paper presented at the Twenty-Second Annual International Pittsburgh Coal Conference on September 13, 2005(1).

The total capital cost only increases by 21 percent when the air-blown system is modified to capture carbon dioxide. However, at the same time the net output drops by 30 percent because of the decreased steam turbine output and the increased auxiliary loads. The combined effects on capital cost per output and COE are dramatic, increasing 72 and 60 percent, respectively.

The steam turbine output decreases because of the large amounts of medium pressure steam needed in the water-gas shift reaction and low pressure steam required to

regenerate the amine. Most of the auxiliary load increase is due to carbon dioxide compression (44.2 MW) and the large amine pump loads (15.8 MW) resulting from using semi-lean amine to reduce the steam regeneration requirement.

Results for oxygen-blown cases with and without CO₂ capture are shown in Table 15. As before, Case 4, rather than Case 2, is chosen for comparison to Case 6 since it uses syngas cleanup rather than stack gas. In Case 2 the contaminants in the coal such as mercury, sulfur and nitrogen were removed from the flue gas after burning the syngas in the combustion turbine. For Case 4 mercury, sulfur and ammonia were removed from the syngas prior to the combustion turbine, similar to Case 6. This makes Case 4 a good reference plant for Case 6. Like the air-blown cases, the total capital cost increases by 21 percent when the system is modified to capture carbon dioxide. The net output decreases for the same reasons as in the air-blown case and by a similar amount, 34 percent. The combined effects on capital cost per output and COE are even more pronounced this time, increasing 84 and 73 percent, respectively.

As discussed above, the oxygen-blown system raises less steam than the air-blown because of the reduced syngas flow through the syngas coolers. But the WGSR and amine system steam demands are similar to those in the air-blown case. So in the oxygen-blown Case 6, so much steam is extracted from the steam cycle that only 19 percent of the steam flow entering the high pressure turbine exits the low pressure turbine. Most of the auxiliary load increase is again due to carbon dioxide compression (43.0 MW) and the large amine pump loads (15.8 MW).

After the Case 5 and 6 modeling and costing efforts were completed, several areas for potential system optimization and cost reduction were identified that may be investigated in follow-up studies. These include changing the coal conveying gas, the amine system configuration, the recycle syngas takeoff point, and the syngas cooling approach. These improvements may have more impact on the oxygen-blown case than on the air-blown, but they are not expected to be significant enough to change the overall conclusions of this paper.

	Case 3	Case 5	% Diff.
Gasifier oxidant	air	air	
CO ₂ capture	no	yes	
Net output, MW	595.5	419.8	-29.5
Net eff. (HHV), %	41.1	28.4	-30.9
ST output, MW	288.3	190.3	-34.0
Auxiliaries, MW	86.8	166.2	91.5
Coal feed, lb/hr	566,400	577,000	1.9
Capital cost, MM\$	763.7	920.3	20.5

Capital cost, \$/kW	1,282	2,192	71.0
O&M, mills/kWh	6.8	9.6	41.2
COE, mills/kWh	40.0	65	62.5

Table 14: Effects of CO₂ removal on air-blown gasification

	Case 4	Case 6	% Diff.
Gasifier oxidant	oxygen	oxygen	
CO ₂ capture	no	yes	
Net power, MW	540.0	344.5	-36.2
Net eff. (HHV), %	41.2	24.9	-39.6
ST power, MW	241.0	135.4	-43.8
Auxiliaries, MW	95.0	174.5	83.7
Coal feed rate, lb/hr	511,500	541,000	5.8
Capital cost, MM\$	813.7	981.8	20.7
Capital cost, \$/kW	1,507	2,850	89.1
O&M, mills/kWh	7.6	11.6	52.6
COE, mills/kWh	44.8	81.7	82.4

Table 15: Effects of CO₂ removal on oxygen-blown gasification

5 CONCLUSIONS

The studies in this report have investigated Transport Gasification systems for power with carbon dioxide capture, comparing oxygen-blown and air-blown gasification.

For these systems using PRB coal, the favored design for producing power with carbon dioxide capture is shown to be air-blown gasification based on lower costs of carbon capture and electricity. It is also shown that the costs of carbon capture from either system are significant.

REFERENCES

- 1 *Power from PRB – Four Conceptual IGCC Plant Designs Using the Transport Gasifier*; Rogers, L. H, et al., Twenty-Second Annual International Pittsburgh Coal Conference; September 13, 2005

- 2 *CO₂ Recovery and Sequestration at Dakota Gasification Company*, M. Perry and D. Eliason, 2004 Gasification Technologies Conference, San Francisco, California
- 3 Dakota Gasification Company carbon dioxide product spec sheet at <http://www.dakotagas.com/Products/Spec%20Sheets/co2spec.pdf>, as of 20-Mar-06
- 4 *MDEA Proven Technology for Gas Treating Systems*, Altofina Chemicals
- 5 *Private Communication*: Ineos Oxide, Freeport TX
- 6 *Gas Purification, 5th Edition*; Kohl A.L. and Nielsen R.B.
- 7 *Engineering Data Book*, Gas Processors Suppliers Association
- 8 *Process Screening Analysis of Alternative Gas Treating and Sulfur Removal for Gasification*, SFA Pacific, Inc., December 2002. Task Order No. 739656-00100.
- 9 “Demonstration of Selective Catalytic Reduction (SCR) Technology for the Control of Nitrogen Oxide (NO_x) Emissions from High-Sulfur Coal-Fired Boilers”, Southern Company Services, Inc., October 1996. DOE Contract DE-FC22-90PC89652.
- 10 *TAG[®] Technical Assessment Guide: Volume 3, Revision 8: Fundamentals and Methods Electricity Supply*, EPRI, Palo Alto, CA, 1999. TR-100281-V3R8.
- 11 *Gasification Technology Status – September 2004*, EPRI, Palo Alto, CA, 2004. 1009769.

DISCLAIMER

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference herein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The view and opinions of authors expressed herein do not necessarily state or reflect those of the United States Government of any agency thereof.

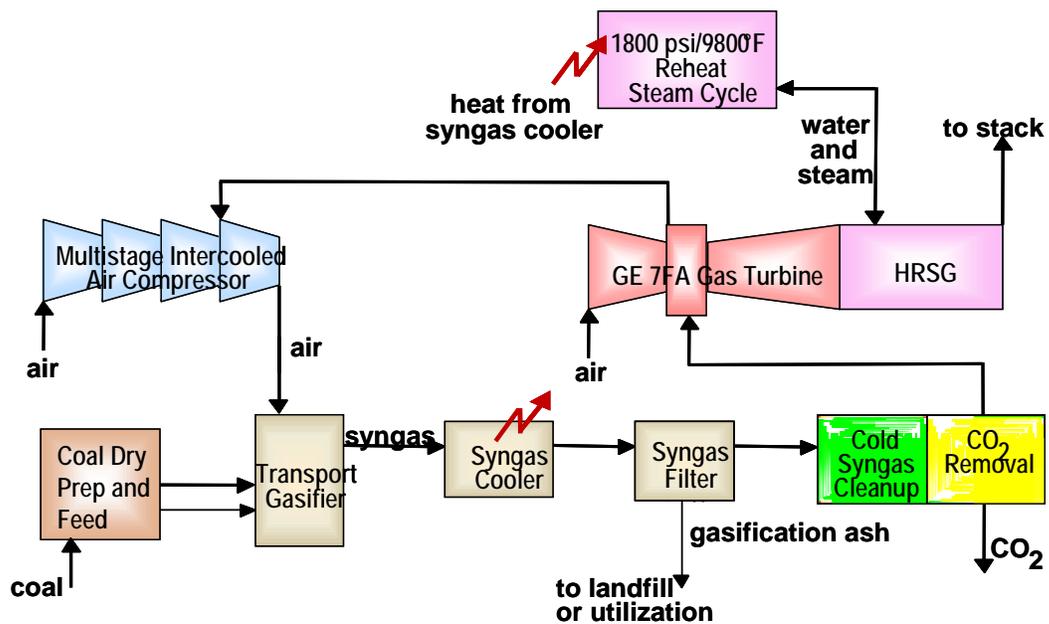


Figure 1: Case 5 simplified process flow diagram

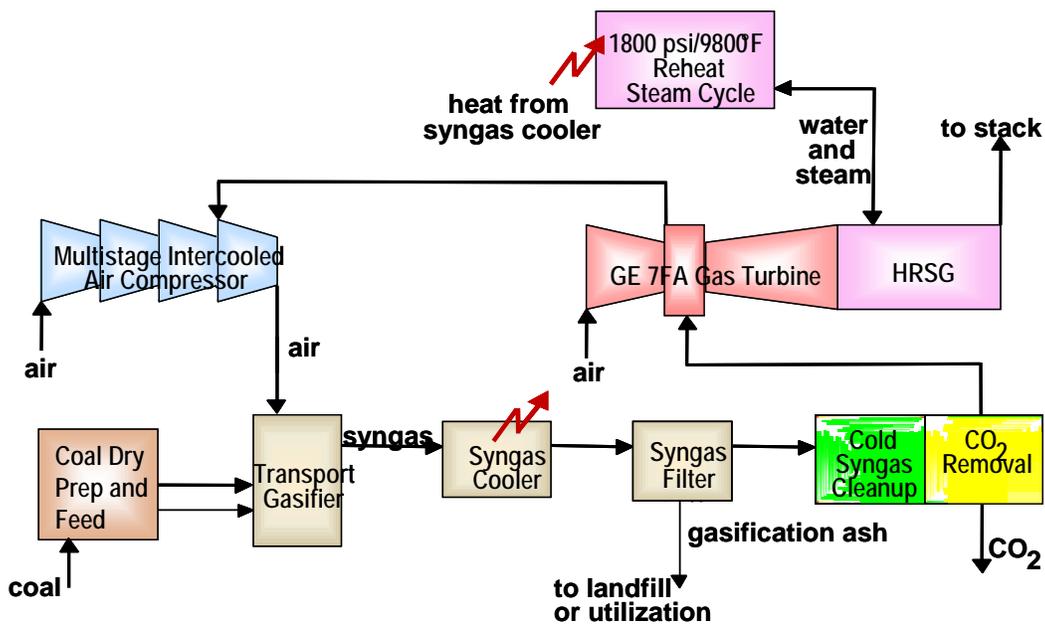


Figure 2: Case 6 simplified process flow diagram

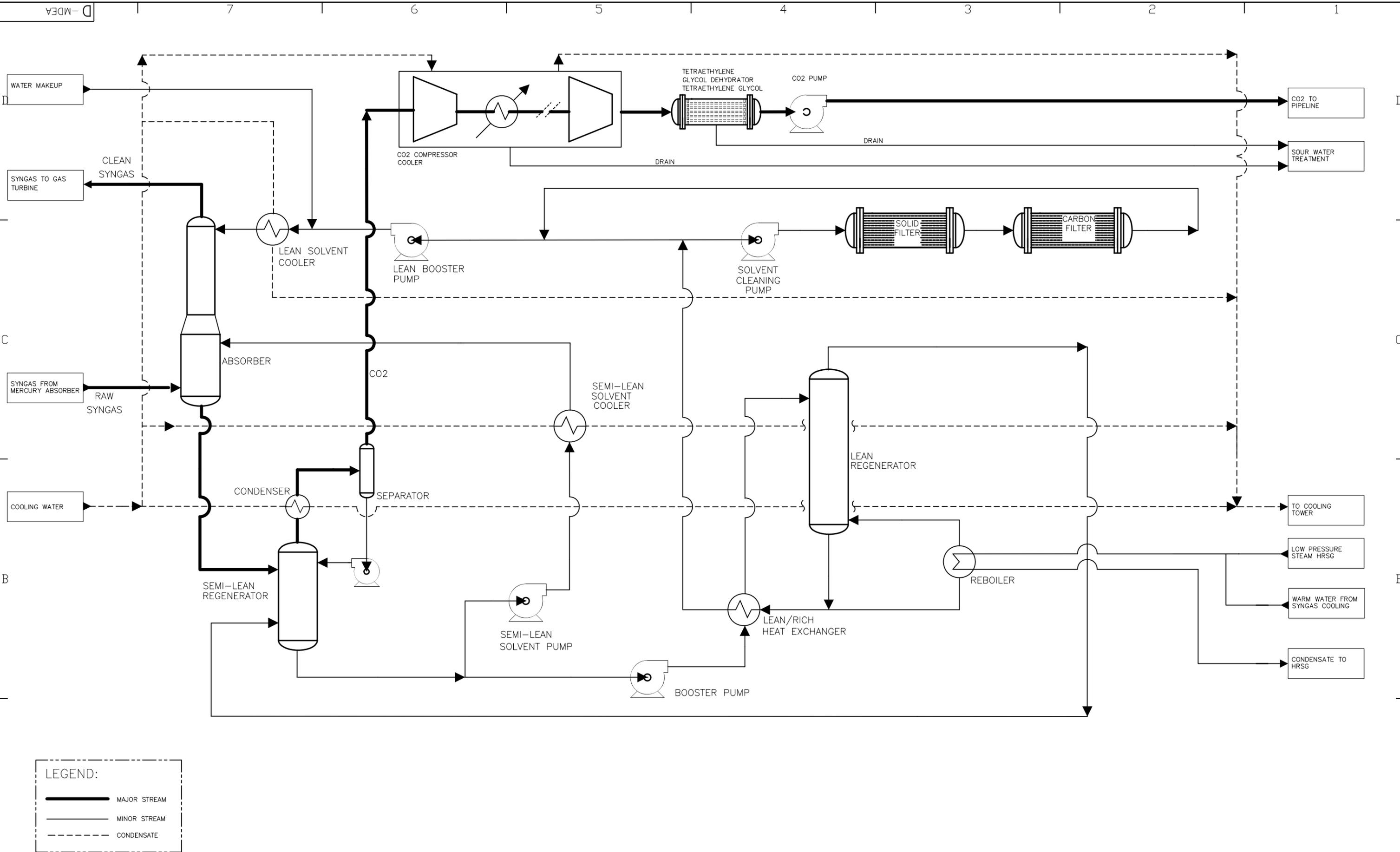


FIGURE 3: Amine process for CO2 capture