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POWER SYSTEMS DEVELOPMENT FACILITY UPDATE ON SIX TRIG™ STUDIES

***Luke H. Rogers*, Principal Engineer, Southern Company Services**
Power Systems Development Facility
31800 Highway 25 N, P.O.B. 1069
Wilsonville, AL 35186
205-670-5989
lhrogers@southernco.com

George S. Booras, Senior Project Manager, Electrical Power Research Institute
P.O.B. 10412
Palo Alto, CA 94303
650-855-2471
gbooras@epri.com

Ronald W. Breault, Project Manager, National Energy Technology Center
U.S. Department of Energy
P.O.B. 880
Morgantown, WV 26507
304-285-4486
ronald.breault@netl.doe.gov

Nicola Salazar, Senior Technology Manager, Coal Monetization, Kellogg Brown and Root, Inc.
P.O.B. 3
Houston, TX 77001
713-753-6833
nicola.salazar@halliburton.com

ABSTRACT

The pursuit of more cost-effective and reliable coal technologies with superior environmental performance continues to drive the development of the Transport Gasifier at the Power Systems Development Facility (PSDF). This paper presents the results of six updated system and economic studies of Transport Integrated Gasification (TRIG™) plants.

Southern Company and Kellogg Brown and Root (KBR), in conjunction with the U.S. Department of Energy (DOE) and other partners, are developing the TRIG™ process at the PSDF for commercial application in the power industry. The PSDF is an engineering scale demonstration of the KBR Transport Gasifier along with a high-temperature, high-pressure syngas filter, a gas cleanup process, and related systems. Built at a sufficient scale to test advanced power systems and components in an integrated fashion, the PSDF provides data necessary for commercial scale-up of these technologies.

To guide future tests and commercialization of the technologies at the PSDF, a series of conceptual commercial plant designs has been completed in partnership with the DOE and the Electric Power Research Institute (EPRI). Six TRIG™ combined cycle cases ha been developed to investigate the relative costs and benefits of oxygen-blown or air-blown gasification, of stack gas or syngas cleanup, and of carbon dioxide capture. These cases are all based on a 2x1 GE7FA+e combined cycle fueled by syngas from two Transport Gasifiers using Powder River Basin (PRB) sub-bituminous coal.

Detailed performance modeling and cost estimation have shown that the optimal configuration for power production without carbon dioxide capture includes air-blown gasification and cold syngas cleanup. Air-blown gasification was also shown to be preferable when carbon dioxide is captured in a TRIG™ combined cycle system.

1. INTRODUCTION

The 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 the capital cost and increase the efficiency of advanced coal-based power generation while meeting strict environmental standards. The PSDF was designed at a scale sufficient to test advanced power system components and Clean Coal Technology Roadmap program elements in an integrated fashion to provide data for commercial scale-up. Clean coal research at the PSDF focuses on testing the Transport Gasifier, a particulate control device (PCD, a high temperature, high pressure gas filter), and a number of auxiliary systems essential for any gasification process.

The Transport Gasifier is an advanced circulating fluidized bed system that offers many advantages over commercially available gasifiers that can lead to successful commercialization. These include high carbon conversion with a variety of fuels, a small footprint with a high thermal throughput, a simple and robust mechanical design, and the ability to easily process high ash, high melting point coals. Syngas produced by the Transport Gasifier can be used as fuel for a combustion gas turbine or a fuel cell, or can be used for processing into chemicals such as methanol or transportation fuels. The Transport Gasifier operates at considerably higher circulation rates, velocities, and riser densities than a conventional circulating bed resulting in higher throughput, better mixing, and higher mass and heat transfer rates. Since the gasifier uses a dry feed and does not slag its ash, it is particularly well-suited for high moisture and ash fuels such as sub-bituminous coal and lignite.

The Transport Gasifier and related systems are being developed for commercial applications by Southern Company, KBR, the DOE National Energy Technology Laboratory (NETL), and other industrial participants, currently including EPRI, Siemens Westinghouse Power Corporation, Peabody Energy, the Lignite Energy Council, and Burlington Northern Santa Fe Railway.

2. COMMERCIAL DESIGN STUDIES

To guide future tests and commercialization of the technologies at the PSDF, a series of conceptual commercial plant designs has been completed in partnership with KBR, DOE NETL, and EPRI. Six Transport Gasification combined cycle cases have been developed to investigate the relative costs and benefits of oxygen-blown versus air-blown gasification and of stack gas or syngas cleanup, as well as the impact of carbon dioxide (CO_2) capture, as shown in Table 1.

	Gasifier Oxidant	Gas Cleanup
Case 1	Air	Stack Gas
Case 2	Oxygen	Stack Gas
Case 3	Air	Syngas
Case 4	Oxygen	Syngas
Case 5	Air	CO_2 Capture
Case 6	Oxygen	CO_2 Capture

Table 1. Six Cases using Transport Gasification

To allow the most meaningful comparisons between cases, a common design basis was used whenever possible. Major similarities include the following:

- Two Transport Gasifiers
- Low-sulfur PRB coal, fed dry
- 2x1 combined cycle plant consisting of 2 GE7FA+e gas turbines and an 1800 psia reheat steam cycle
- Dry ash removal from the syngas at 500 – 700°F by metal filter elements
- Selective Catalytic Reduction (SCR) in the Heat Recovery Steam Generator (HRSG) to control nitrogen oxides (NO_X)
- 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 were selected for these configurations that are either commercially available or are available with commercial guarantees in the 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 six system designs that are each 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 or sequestration-ready carbon) the systems would be configured differently and the results would be different.

The performance and cost for each case have been calculated 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. Detailed operation and maintenance plans, reliability and maintainability models, and emissions calculations have also been developed.

The assumptions and calculations in these studies are representative of an Nth plant. For the configurations selected, the technologies are assumed to be mature.

3. COMMERCIAL SYSTEM DESCRIPTIONS

Since the six cases have common features, they will be described together. Differences will be noted in the text. For simplicity, a single train of equipment will be described in the text, even though several systems, such as the coal feed systems, are spared. Simplified process flow diagrams for the six cases are given in Figures 1-3.

Each case consists of two identical gasification trains, each consisting of a single gasifier fueling one gas turbine which exhausts into a dedicated HRSG. The output from the two HRSGs is combined in a single steam cycle.

3.1 Solids Handling, Preparation, and Feed Systems

The design fuel is sub-bituminous Powder River Basin coal with an as-received higher heating value of 8,760 Btu/lb, 27 percent moisture, and 0.26 percent sulfur. Coal preparation consists of coal unloading, coal reclaim and crushing, coal drying and milling, and high pressure coal feed system.

Roll mill pulverizers incorporate a flash dryer in which hot gas dries the coal to approximately 18 percent moisture. Dusty gas generated from the pulverizers passes through a baghouse to remove fines before being cooled to condense out the moisture in a knockout drum. The gas from the knockout drum is reheated via medium pressure steam and recirculated back to the pulverizer to dry more coal.

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 1, 3, and 5 and by recycle syngas in Cases 2, 4, and 6 via dilute-phase conveyors. The PRB coal composition as fed to the gasifier is shown in Table 2. The as-fed coal higher heating value is 9,850 Btu/lb.

	Weight %
C	58.4
H	3.8
O	13.0
N	0.8
S	0.3
Ash	5.8
H_2O	17.9
Cl	0.012
F	0.004
Hg	8×10^{-6}

Table 2: Coal properties, as fed to the gasifier

3.2 Gasification

The Transport Gasifier operates as a pressurized, circulating fast-fluidized bed gasifier and consists of simple refractory-lined pipe sections. The operating temperature range is approximately 1,700 to 1,800°F in these

studies, and the operating pressure is just over 400 psia for all cases, with the cases incorporating syngas cleanup being somewhat higher.
Low-sulfur PRB sub-bituminous coal, an oxidant, and steam are converted into a particulate-laden syngas in the gasifier with a carbon conversion of 97 percent. Particulate-free syngas and/or off gases are recycled to the gasifier for solids fluidization. During system start-up, natural gas-fired burners heat the gasifier before solids are introduced. A startup/makeup solids feed system adds material to the gasifier when needed.

In the air-blown configurations (Cases 1, 3 and 5) a multi-stage, intercooled, motor-driven process air compressor supplies most of the air required by the gasifier, and the balance is extracted from the gas turbine. This arrangement has two benefits: it allows the power output of the gas turbine to be maximized at different ambient conditions by varying the relative air flow rates, and it increases the operational flexibility of the system. The air extracted from the gas turbine compressor is cooled in the process air recuperator and then added inter-stage to the process air compressor. The combined flow is further compressed, and a small stream is diverted to convey solids to the gasifier. The rest of the air is heated in the recuperator and then sent to the gasifier. A small nitrogen plant provides purge gas to the plant.

In the oxygen-blown cases (Cases 2, 4 and 6) a conventional cryogenic Air Separation Unit (ASU) provides the gasifier oxidant, purge nitrogen, and off gases. An air extraction is taken from the combustion turbine (CT) in Cases 2 and 4 and combined with air from the ASU main compressor. The major off gas stream from the ASU (mostly nitrogen) is compressed and sent to the CT to maximize output and minimize NO_X formation. In Case 6 this stream is humidified to make up for the mass lost in the CO_2 that is removed from the syngas. The 95 percent pure oxygen from the ASU is sent to the gasifier via an intercooled multi-stage compressor.

In the configurations that use stack gas cleanup (Cases 1 and 2), dry limestone is fed to the gasifier for in-situ sulfur capture during the gasification process. This lowers the sulfur level entering the flue gas treatment system, decreases ammonium bisulfate formation in the HRSG, and lowers the acid dew point temperature in the HRSG, allowing more heat to be recovered. However, it also increases the amount of gasification ash that must be captured, handled, and disposed.

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 at 500 - 700°F in the PCD. Based on PSDF experience, a minimum filter operating life of 8,000 hours is expected. A safeguard device (failsafe) is installed in each filter element to protect the gas turbine from particulate-related damage in the event of a filter element failure. In the configurations using stack gas cleanup (Cases 1 and 2), the syngas is sent directly from the syngas filter to the CT at 700°F, without further cooling or treatment.

The gasification ash consists of un-reacted carbon and coal ash, plus, in Cases 1 and 2, reacted and un-reacted limestone. It is cooled and depressurized and then sent to storage silos, from which it is removed by truck to a 5-year, dry storage landfill. Tests at the PSDF have shown that the material is non-hazardous and that it can be handled like conventional coal combustion ash.

3.3 Case 3 and 4 Syngas Treatment

In Cases 3 and 4, the particulate-free syngas leaving the PCD is cooled in the syngas recuperator before it is passed through a reactor where carbonyl sulfide (COS) is hydrolyzed to hydrogen sulfide (H_2S). The catalyst for the reaction is alumina-based and is chlorine-resistant. Then the syngas is further cooled to 145 - 150°F by heating cleaned syngas and by warming steam cycle condensate, recovering the heat into the steam cycle. Next, mercury is removed from the syngas when it passes through a packed bed of sulfur-impregnated activated carbon.

The syngas then goes to the CrystaSulf® H_2S absorber, where almost all of the H_2S present in the syngas is absorbed by a proprietary mixture of organic solvents saturated with sulfur dioxide (SO_2). The SO_2 oxidizes the absorbed H_2S to produce elemental sulfur. The proprietary chemical holds the elemental sulfur in solution until it is removed in separate vessels by crystallization and filtration. The CrystaSulf process is ideally suited for treating low sulfur syngas streams: it does not adsorb CO_2 and it has lower capital and operating costs than other commonly used systems. The sulfur is discharged from the CrystaSulf process as a nearly-pure sulfur cake, but no byproduct credit is taken in the financial calculations of these studies. After the CrystaSulf process the clean syngas is reheated to 350°F in the syngas recuperator and then sent to the gas turbine.

Sour water condenses from the syngas as it is cooled below the dew point. The water dissolves almost all the nitrogenous compounds, chloride, and fluoride present as well as lesser amounts of H_2S, COS, CO, and CO_2. 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 recycled to the process through compressors. 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 fluidization, filter element back pulsing, and–in the oxygen-blown cases–coal conveying.

3.4 Case 5 and 6 Syngas Cleanup and Carbon Dioxide Removal

In Cases 5 and 6, which incorporate CO_2 removal from the syngas, the syngas is fed after the PCD to fixed-bed catalytic reactors to convert most of the CO to CO_2 and H_2. When sub-bituminous coal is gasified in the Transport Gasifier, 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. This CO level must be greatly reduced before combustion to achieve a high overall carbon capture. The reaction of CO with steam to produce hydrogen, the water gas shift reaction (WGSR), is exothermic. This results in a loss of heating value, so the coal feed system, gasifiers, ASU, etc. are larger for the CO_2 capture cases than for the non-capture cases.

Since the syngas in Cases 5 and 6 contains sulfur compounds, the catalysts for a typical two-step WGSR approach can not be used. Instead, a cobalt-molybdenum catalyst is used that operates at 550 ºF and is not susceptible to reduced sulfur compounds poisoning. In fact, this catalyst requires about 100 ppmv of H_2S in the syngas for it to be effective and is thus referred to as a "sour shift" catalyst. The WGSR catalyst also catalyzes the hydrolysis of HCN in the syngas to NH_3 and of COS to H_2S and CO_2, thus enabling nearly all the sulfur to be captured as H_2S by the downstream amine-based CO_2 removal system.

Like all WGSR catalysts, the cobalt-molybdenum catalyst is poisoned by particulates, chlorides and condensed water. The PCD 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 water vapor in the syngas is not sufficient to achieve a high conversion of CO, so extra steam is added. The steam consumption is reduced by splitting the WGSR process into three reactors in series and optimizing the conversion in each reactor. The catalyst temperature limit is 900°F, so the syngas is cooled to 550°F before the second and third reactors.

After the WGSR, the syngas is cooled to 100°F, and mercury is removed from the syngas in a bed of sulfur-impregnated activated carbon. Sour water is condensed out of the syngas and processed in the same manner as described above for Cases 3 and 4.

The syngas is then sent to an amine-based scrubbing system to absorb CO_2 (and other acid gases such as H_2S). To reduce the solvent regeneration energy cost and to ensure high CO_2 removal efficiency, a split stream scrubbing system incorporating lean and semi-lean amine solution streams is used. The raw syngas enters the bottom of the absorber where the bulk of the CO_2 is scrubbed with nearly 80 percent of the semi-lean solution, which is 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 CO_2 is stripped to produce lean amine solution. It is fed to the top of the absorber as a polishing agent to complete the scrubbing of the CO_2 from the syngas. The rich amine solution from the bottom of the CO_2 absorber expands from the high absorber pressure to the lower pressure of the semi-lean regenerator through a hydraulic turbine, which provides some of the power required to drive the semi-lean solution pumps.

In the lean regenerator, the CO_2 and other acid gases such as H_2S 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 ^oF. 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 CO_2 from the rich amine exiting the bottom of the absorber. The final vapor and acid gases leaving the semi-lean regenerator are cooled by a condenser, then fed to a separator to knock out the moisture, and then recycled to the semi-lean regenerator. The acid gas leaving the top of the separator is mainly CO_2 that is ready for recompression and storage or transportation. The absorber is operated at the pressure of the syngas to favor the CO_2 absorption while the semi-lean regenerator and the lean regenerator are operated at a much reduced pressure (25 psia) to enhance the CO_2 stripping.

The carbon dioxide-rich stream produced is dried and compressed to pipeline pressure (2,200 psia) by the CO_2 compression system. The system includes a 4 stage, intercooled compressor, a TEG dehydration package, and a multi-stage pump.

As mentioned above, 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 (Reference 1). The carbon dioxide sold to PanCanadian Petroleum Limited for enhanced oil recovery in Saskatchewan, Canada, contains 1 percent H_2S, by volume (Reference 2). This cost-saving approach is only acceptable if the CO_2 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.

3.5 Combined Cycle Island

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 and replacing the stage one turbine nozzle to accommodate the increased mass flow associated with the dilute syngas 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 the oxygen-blown cases) 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 syngas composition entering the gas turbine in the six cases is given in Table 3. Note that the syngas in Cases 1 and 2 below has not undergone cleanup, but the pollutants are removed post-combustion in the stack gas treatment system described in Section 3.6.

	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
CH_4	2.10	2.50	2.21	3.07	2.39	3.70
CO	22.59	34.30	23.74	42.14	1.24	2.37
CO_2	6.76	13.58	7.03	16.59	0.03	0.01
H_2	11.47	29.20	12.05	35.86	37.56	91.80
HCN	0.016	0.032	0.008	0.013	0	0
H_2O	5.58	18.89	1.02	0.90	0.20	0.40
H_2S	0.012	0.056	0.001	0.001	0	0
N_2	51.32	1.16	53.93	1.42	58.58	1.72
NH_3	0.152	0.280	0.007	0.001	0	0

Table 3. Syngas Compositions Entering the Gas Turbine, by Volume %

The HRSG is a dual-pressure unit with single reheat. 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 feed water pump outlets as 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, except in Case 3, in which excess medium pressure steam generated in the gasification island is added to the HRSG cold reheat flow. The resulting cold reheat steam is reheated in the HRSG and then expanded in the medium pressure section of the steam turbine.

In the oxygen-blown cases low pressure steam is taken from the crossover as the steam is sent to the low pressure steam turbine and sent to the ASU. In Cases 3 - 6, low pressure steam from the same source is sent to the reboilers in the sour water treatment system. In Cases 5 and 6 a large flow of 65 psia steam is extracted from the steam turbine for use in the amine system. The exhaust from the low pressure turbine is condensed at 1 psia by water from a wet mechanical draft cooling tower, and then sent to the gasification island for use in process cooling. The cooling tower also supplies cooling water for other process areas as needed, such as process air compressor intercoolers, ASU, coal mill systems, limestone mill systems, and the CrystaSulf system.

The HRSG exhaust temperature is maintained at least 20°F above the acid dew point temperature of the flue gas in each case. Although the HRSG stack temperatures are quite different between the cases, the pinches used in the HRSG are similar. The only exception is that larger pinches are used in Cases 1 and 2 because the sulfur remaining the the CT exhaust necessitates higher stack temperatures. A SCR system is installed in the HRSG to reduce NO_X emissions. In Cases 3 - 6, 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.

3.6 Stack Gas Treatment

In Cases 1 and 2, almost all of the sulfur is removed from the HRSG exit flow in the stack gas treatment system, which is based on the Chiyoda CT-121™ flue gas desulfurization process. The gas turbine exhaust leaving the HRSG is cooled and saturated in a spray chamber, and then passes through a limestone slurry which removes the sulfur, as well as most of the chlorides and fluorides in the gas. A motor-driven fan provides the pressure difference to drive the stack gas through the stack gas treatment system.

4. 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 were modeled using Aspen Plus® from AspenTech, Inc. Gas cleanup performance and costs for the CrystaSulf systems were provided by CrystaTech, Inc., based on data from their pilot plant. KBR provided the composition of the sour water condensed from the syngas and data for the sour water treatment systems. The combined cycle in each Case was modeled using Thermoflex from Thermoflow, Inc, with the gas turbine model calibrated using 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 data.

Gasification-related equipment, solids handling equipment, and balance of plant equipment were sized using Southern Company-propriety routines, by using Kbase software from AspenTech, or were sized by equipment vendors. Combined cycle equipment was sized using Thermoflex or GT PRO software, both from Thermoflow, Inc. Equipment costs were 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. These cases were developed over several years, so 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% feedwater pumps and 7 x 17% coal mill systems.

Operations and Maintenance (O&M) cost estimates for each case were developed from the component-level calculations rather than by applying rule-of-thumb factors to the overall capital cost of the plant. For each case, detailed O&M models were 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-fired and natural gas-fired power plants, and KBR's experience with the operation of their fluid catalytic cracking units.

A Reliability, Availability, and Maintainability (RAM) analysis has been completed on Cases 1 and 2 using UNIRAM software from EPRI, using the full equipment lists and sparing philosophy of both cases. As both the RAM analysis and the O&M cost model start at the component level, much of the source data for reliability and repair costs is taken from general industrial sources, since most equipment used in these systems has decades of history in other applications.

5. RESULTS

The results of these studies are summarized below. Although emissions rates were calculated, they are not presented in these results.

5.1 Performance

The performance of the six configurations is given in Table 4.

	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Net Power Output, MW	573.9	511.9	594.6	530.9	428.0	361.0
Net Efficiency (HHV), %	42.1	39.9	41.0	40.0	29.0	26.1
Net Efficiency (LHV), %	44.6	42.2	43.4	42.3	30.7	27.6
Heat Rate (HHV), Btu/kWh	8,100	8,560	8,320	8,530	11,800	13,100
Heat Rate (LHV), Btu/kWh	7,650	8,090	7,860	8,060	11,100	12,400
Coal Feed Rate, lb/hr	532,000	502,000	566,000	519,000	577,000	541,000
Gas Turbine Output, MW	394.0	394.0	394.0	394.0	394.0	394.0
Steam Turbine Output, MW	266.1	209.2	287.4	230.1	196.8	132.5
Plant Auxiliary, MW	86.2	91.3	86.8	93.2	162.8	165.5

Table 4. Power, Efficiency, and Heat Rate

Note that in each case the syngas flow is the amount necessary to fire the two gas turbines to their full output. Therefore, the coal feed rates are different.

The air blown systems (Case 1 ,3 and 5) produce more power than the oxygen-blown systems (Case 2, 4 and 6) primarily because the larger syngas flow raises more high pressure steam in the syngas cooler. Also, the power consumed by the process air compressor in the air blown cases is less than that consumed by the ASU in the corresponding oxygen-blown cases. The coal feed rate is higher in the air-blown cases because of the coal required to heat the nitrogen in the air to the gasifier outlet temperature.

The systems with syngas cleanup (Cases 3 and 4) have a higher net output than the systems with stack gas cleanup (Cases 1 and 2) because of the absence of stack gas cleanup system fans and because the cooler syngas temperature to the CT necessitates a higher syngas flow and thus more steam is generated in syngas cooling.

Comparing the carbon capture cases (5 and 6) with their non-capture counterparts (Cases 3 and 4 are the most similar) shows a significant impact on output and efficiency. 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 solution. Most of the auxiliary load increase is due to carbon dioxide compression (44.2 MW) and the semi-lean amine pumps (9.4 MW).

5.2 Capital Costs

The capital costs are reported in Table 5 in January, 2003, U.S. dollars.

	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Indirects	250.4	259.3	252.1	261.2	290.0	300.0
Equipment Costs	502.3	541.2	511.6	552.5	630.3	681.8
TOTAL, MM$	752.7	800.5	763.7	813.7	920.3	981.8
TOTAL, $/kW	1,312	1,564	1,284	1,533	2,150	2,720

Table 5. Capital Costs in Million Dollars, January, 2003

The Indirects category includes expenses such as AFUDC, engineering, project management, construction management, temporary facilities and services during construction, startup, insurance, taxes, and land. Recent escalations in commodity prices such as steel and concrete since January, 2003, are not reflected in the above capital costs.

The total capital costs for the oxygen-blown cases are roughly $50 million higher than that of the air-blown counterparts, giving the air-blown systems significantly lower costs per kilowatt. The cost savings in the oxygen-blown systems from smaller gasifiers, syngas coolers, syngas filters, etc. is more than offset by the cost savings in the air-blown system of using an air compressor rather than an ASU.

The total cost differences are small between the configurations with stack gas cleanup (Cases 1 and 2) and syngas cleanup (Cases 3 and 4). This shows that for these designs, the cost of syngas cleanup is not prohibitively high, even for air-blown gasification.

The added cost of the carbon capture-related equipment is similar for both the air-blown and oxygen-blown cases (comparing Case 5 with 3 and Case 6 with 4). Because the amount of carbon dioxide captured is similar in both the air-blown and oxygen-blown cases, the size and cost of the amine systems are similar. Even the absorber vessels are approximately the same size in both cases, despite the widely different syngas flows, since they are sized on the amine flow rather than the syngas flow.

5.3 Operating and Maintenance Costs

The O&M costs are reported in Table 6, distributed between Fixed and Variable O&M according to the EPRI TAG® basis (Reference 3). This basis shifts many costs that might normally be considered variable O&M to fixed O&M. 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. Variable O&M includes the costs of expendables and natural gas for startup and flare.

	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Fixed O&M, $/kW-yr	42.30	48.40	43.20	48.80	57.80	71.50
Variable O&M, $/MWh	0.84	0.97	0.59	0.62	1.14	1.24

Table 6. Operating and Maintenance Costs

The variable O&M costs are lower for the syngas cleanup configurations (Case 3 and 4) than for the stack gas cleanup configurations (Case 1 and 2) because of the ammonia produced in the sour water treatment system. Part of the ammonia produced is used in the SCR, eliminating the need to purchase ammonia, and a byproduct credit is taken for the rest at 75 percent of the market value. Without this byproduct credit, the variable O&M for Cases 3 and 4 would be higher than for Cases 1 and 2 because of the yearly purchase of the CrystaSulf

solution. The higher O&M costs for Cases 5 and 6 are primarily due to the WGS and amine systems, and also to the lower net output. The oxygen-blown cases show higher O&M costs than the corresponding air-blown cases because of ASU maintenance and lower net output.

5.4 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 (COE) for each case are reported in Table 7, in January, 2003 dollars.

	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Capital, mills/kWh	23.4	27.9	22.9	27.3	39.6	50.5
O&M, mills/kWh	6.9	7.9	6.8	7.6	9.4	11.4
Fuel, mills/kWh	10.1	10.7	10.4	10.7	14.8	16.4
Total COE, mills/kWh	40.4	46.4	40.0	45.6	63.7	78.3

Table 7. Levelized Capital, O&M, Fuel, and Electricity Costs

The values in Table 7 do not all sum to the given totals due to rounding errors. The calculated values for AFUDC that are included in the costs in Table 5 are subtracted out of the total costs for the COE calculations so that the standard carrying charge factor can be used.

The costs of electricity above show that the optimal TRIG™ configuration for power production without carbon dioxide capture includes air-blown gasification and syngas cleanup.

5.5 Cost of Carbon Capture

The costs of carbon dioxide capture and avoidance for Cases 5 and 6 are shown in Table 8.

	Case 5	Case 6
Gasifier oxidant	air	oxygen
CO_2 capture cost, $/ton	21.3	25.9
CO_2 avoided cost, $/ton	32.4	41.8

Table 8. Costs of Carbon Capture

Since the effects of CO_2 capture on performance and cost are fairly independent of the gasifier oxidant in these cases, the consistent air-blown system COE advantage carries over into a cost of carbon capture advantage for the air-blown Case 5.

The cost of electricity advantage for air-blown gasification shown in Table 7 and the costs of carbon capture advantage shown in Table 8 make it clear that for these TRIG™ configuration with amine-based carbon capture, air-blown gasification is preferable.

To achieve an overall carbon capture of approximately 90 percent in these cases, the amine systems are pushed to nearly complete carbon dioxide capture. It is expected that further studies with only 80 - 85 percent carbon capture will result in significantly lower costs and better performance, more in line with the increases shown for other IGCC systems (Reference 4).

These two studies of Case 5 and 6 were the PSDF's first detailed look at applying carbon capture to the TRIG™ process, so they should be viewed as a starting point, not as the final word. Evaluations of better configurations and more economic carbon capture technologies are on-going.

5.6 Reliability, Availability and Maintainability

A Reliability, Availability and Maintainability (RAM) analysis was conducted on Cases 1 and 2 to calculate the overall yearly availability and to determine which pieces of equipment have the greatest impact on RAM. A liquid oxygen (LOX) system was included in Case 2 as a backup to the ASU, sized to make the oxygen supply in Case 2 as reliable as the air supply in Case 1. Thus the two cases have approximately the same overall Equivalent Availability Factor (EAF), which is calculated to be 89.7 percent with a CT backup fuel, and 83.7 percent with no CT backup fuel. The Effective Forced Outage Rate (EFOR) with backup fuel is approximately 6.5 percent for both cases. These calculations are based on a twelve year maintenance cycle with 7.6 percent unavailability due to planned outages.

This projected availability is higher than the 75 to 80 percent currently achieved by oxygen-blown, entrained-flow IGCC plants (Reference 5). This is primarily because the Transport Gasifier operates at a much lower temperature than an oxygen-blown, entrained-flow gasifier, extending refractory life and reducing the number and duration of outages, and because the Transport Gasifier uses dry feed systems rather than burner nozzles that require frequent replacement in slurry-fed, oxygen-blown IGCC plants.

The pieces of equipment with the greatest impacts on RAM are listed below in order of impact on EAF. All other pieces of equipment account for less than 3 percent of the down time.

1. Gas Turbines
2. Steam Turbine
3. Syngas filters
4. HRSGs
5. ASU (for oxygen-blown, if backup LOX were not included)
6. Secondary Syngas Coolers
7. Primary Syngas Coolers
8. Transport Gasifiers

Because of the simple design of the Transport Gasifier, its impact on unavailability is low.

6. CONCLUSIONS

The results of these studies have shown that for the production of power from low-sulfur PRB coal with currently available technologies, air-blown Transport Gasification is preferable to oxygen-blown Transport Gasification. They have also demonstrated that for these project configurations and design assumptions, syngas cleanup is preferable to stack gas cleanup. It has also been shown that the costs of carbon capture from either the oxygen blown or air blown systems are significant, but the air-blown system still gives lower overall costs.

7. FUTURE RESEARCH AND STUDY

The PSDF is currently undergoing the first test campaign since a number of gasifier modifications were made to improve carbon conversion and increase the syngas LHV. Early test results show that the modifications were very successful. Future studies will include the improved Transport Gasifier performance.

The Transport Gasifier is the foundation of an IGCC unit in Orlando, Florida, that is part of DOE's CCPI program. Southern Company and KBR are currently in the design phase of the project, and the projected start date is mid-2010. Information gained during the design of the Orlando Gasification Project will be incorporated in future commercial design studies.

Subjects to be investigated in future commercial design studies may include carbon capture at lower levels, the use of other gasifier fuels, and the incorporation of more advanced gas turbines.

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