Process Engineering Division

Transport Gasifier IGCC Base Cases

PED-IGCC-98-006

September 1998

Latest Revision June 2000

PREFACE

This report presents the results of an analysis of two Transport Gasifier IGCC Base Cases. The analyses were performed by W. Shelton and J. Lyons of EG&G.

EXECUTIVE SUMMARY

TRANSPORT HGCU GASIFIER IGCC BASE CASES

EXECUTIVE SUMMARY

ASPEN PLUS (version 10.1) Simulation Models and the Cost of Electricity (COE) have been developed for two IGCC cases based on the Transport gasification process. The objective was to establish base cases for commercially available (or nearly available) power plant systems having a nominal size of 400 megawatts (MWe). The simulation models are based on previous simulations (ASPEN Archive CMS Library), available literature information, and published reports. The COE estimates were based on data from the EG&G Cost Estimating Notebook and several contractor reports. These cases can be used as starting points for the development and analysis of proposed advanced power systems.

Both cases include the following process sections:

- Transport Gasifier with in-bed sulfur capture.
- Sulfator with the exhaust gas combined with the gas turbine exhaust.
- Chloride guard reactor and a Transport Desulfurizer hot gas cleanup section.
- “G” gas turbine -W501G modified for coal derived fuel gas.
- Three pressure level subcritical reheat Steam Cycle
  - (1800 psia/1050°F/342 psia/1050°F/ 35 psia).

The oxidant used in the gasifier accounts for the major difference between the cases. Case 1 is an air-blown gasifier system while Case 2 is an oxygen-blown system. The raw fuel gas cooler section following the gasifier (and integrated with the gasifier and other heat exchangers) is used for generating high-pressure superheated steam. This section is followed by a ceramic filter that captures particulates for recycle to the gasifier and a chloride guard bed. The fuel gas then enters a hot gas cleanup unit (HGCU) using a transport absorber/regenerator process. The sulfur dioxide rich waste stream from the HGCU is sent to a sulfator. Power is recovered for both cases using a modified W501G gas turbine and a three-pressure level reheat steam cycle.

Process flow diagrams and material and energy balances summaries are shown in Figures 1 - 4 and COE summaries are given in Appendix A. In Table 1 the overall results obtained for power generation, process efficiency, and COE are compared for the cases.

Table 1: Transport Gasifier IGCC Base Cases Summary

	CASE 1	**CASE 2**
Gasifier	Transport Air-Blown	Transport Oxygen-Blown
Sulfur Removal	In-Bed Sulfur Capture/HGCU	In-Bed Sulfur Capture/HGCU
Gas Turbine Power (MWe)	272.6	272.6
Steam Turbine Power (MWe)	162.6	142.4
Misc./Ax Power (MWe)	-20.0	-31.3
Total Plant Power (MWe)	415.4	383.7
Efficiency, HHV (%)	49.8	47.1
Efficiency, LHV (%)	51.7	48.8
Total Capital Requirement, ($1000)	484,062	496,722
$/kW	1165	1295
Net Operating Costs ($1000)	45,388	47,294
COE (mills/kWh)	38.1	41.9

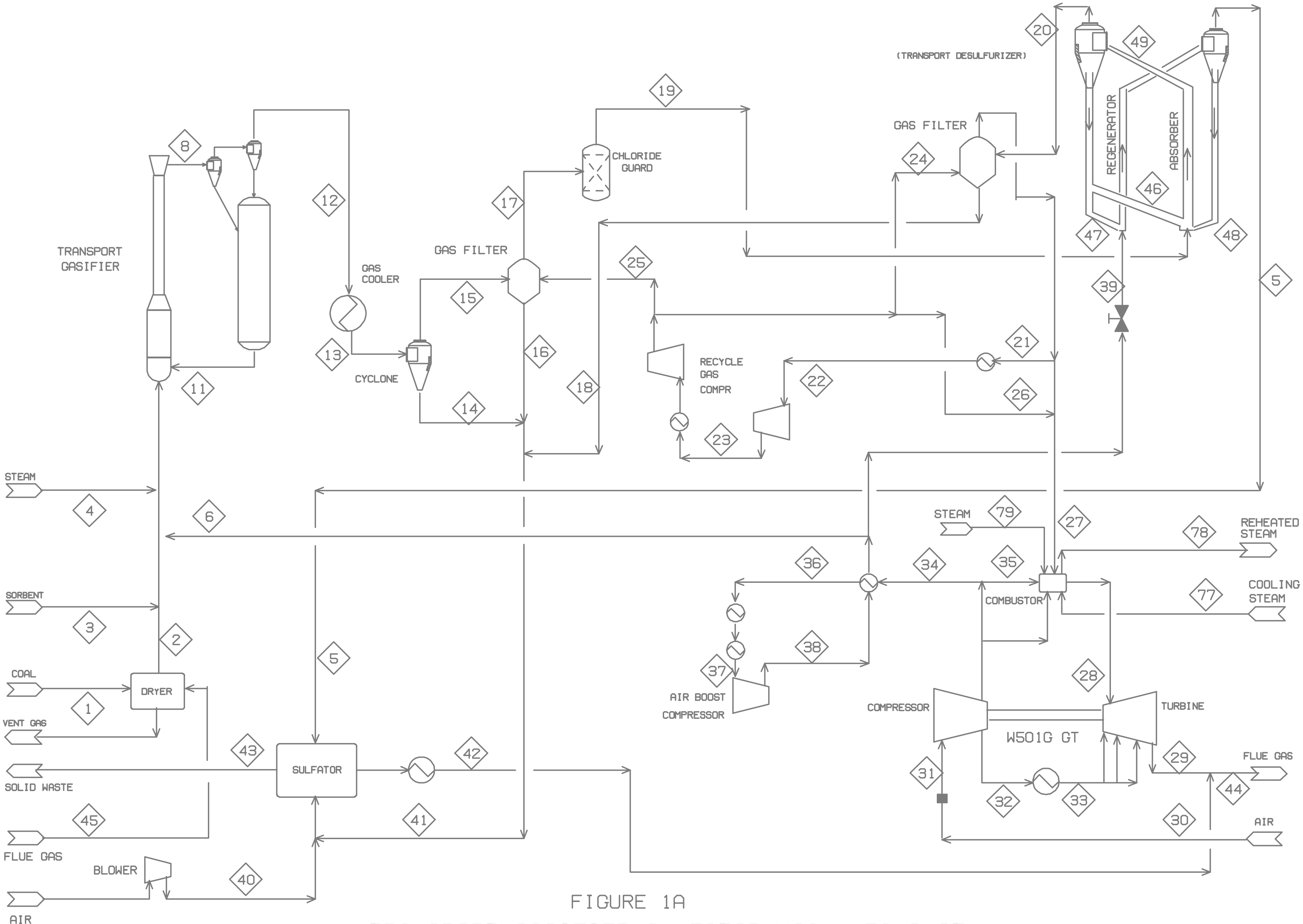

FIGURE 1A

TRANSPORT GASIFIER/SULFATOR/HGCU/W501G GT

(85% in-bed Sulfur Capture)

FIGURE 1B

TRANSPORT GASIFIER / SULFATOR / HGCU / W501G GT / 3 PRES LEVEL STEAM CYCLE

SUMMARY:

POWER	MWe	EFFICIENCY	%
GAS TURBINE	272.8	HHV	49.8
STEAM TURBINE	162.6	LHV	51.7
MISCELLANEOUS	7.2		
AUXILIARY	12.8		
NET POWER	415.4		

STREAM	1	2	3	4	5	6	8	11	12	13	14	15	16	17
FLOW (LB/HR)	243937	228222	30634	77596	6484	657280	5775676	4781947	994045	994045	45884	948161	2294	954722
TEMPERATURE (F)	59	200	200	630	1338.2	650	1657.9	1647.9	1657.9	1004	1004	1004	1000	1000
PRESSURE (PSIA)	14.7	14.7	14.7	400	356	400	395.2	400	395.2	385.2	367	367	362	362
H (MM BTU/HR)	303.8	406.8	-159.2	-429.8	-1.8	67.2	-4864.1	-4224.8	-664.6	-895	-53.4	-841.7	-2.7	-848.3

STREAM	18	19	20	21	22	23	24	25	26	27	28	29	30	31
FLOW (LB/HR)	121	954042	953588	14367	14367	14367	4310	8907	1149	944560	4084656	4611765	4320000	4320000
TEMPERATURE (F)	1047.9	997.6	1050	1047.9	450	490.3	565.3	565.3	565.3	1047.3	2584.2	1128	59	59
PRESSURE (PSIA)	337	357	347	337	327	375	750	750	750	336	268.5	15.2	14.7	14.6
H (MM BTU/HR)	-0.1	-848.4	-852.3	-12.9	-15.7	-15.5	-4.5	-9.4	-1.2	-845	-565.2	-2273	-186.5	-186.5

STREAM	32	33	34	35	36	37	38	39	40	41	42	43	44	45
FLOW (LB/HR)	527109	527109	663316	3116096	663316	663316	663316	6036	111073	48299	110052	55803	4721818	555540
TEMPERATURE (F)	809.7	600	809.7	809.7	365.8	128	203	640	161.1	1005.1	1200	300	1129.7	465.8
PRESSURE (PSIA)	282.2	276.6	282.2	282.2	281.2	275	405	362	25	337	16	14.7	15.2	15.2
H (MM BTU/HR)	76.3	47.7	96	450.9	20.8	-19.8	-6.6	0.6	-2.1	-56.2	-42.3	-145.7	-2315.3	-373

STREAM	46	47	48	49	77	78	79
FLOW (LB/HR)	649300	72144	71697	1675038	70000	70000	24000
TEMPERATURE (F)	1030	1030	1338.2	1050	607.3	1050	602
PRESSURE (PSIA)	347	347	357	352	350	342	350
H (MM BTU/HR)	-2233.1	-248.1	-248.7	-3330.2	-388.6	-372	-133.2

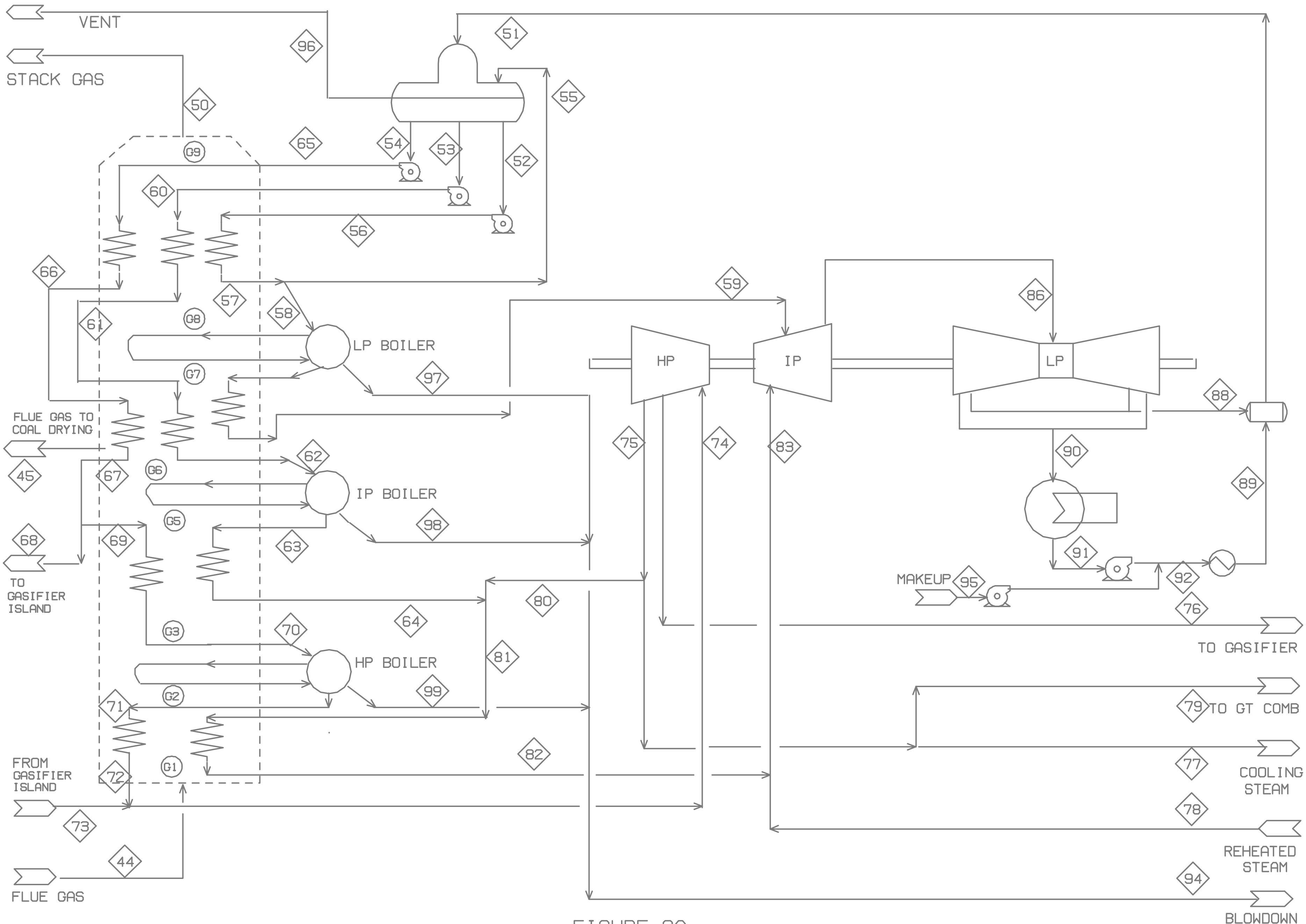

FIGURE 2A
TRANSPORT GASIFIER IGCC – STEAM CYCLE

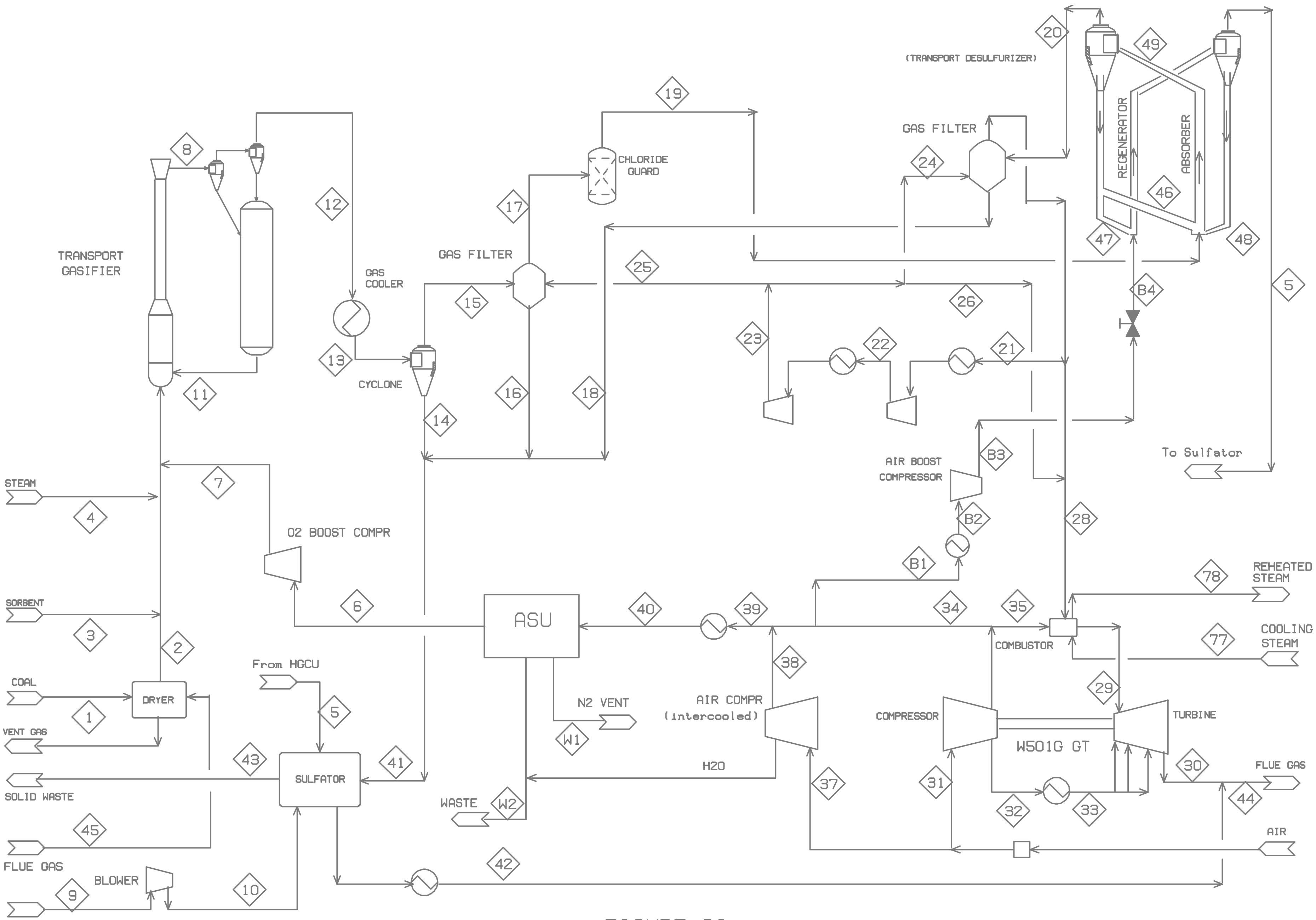

FIGURE 3A
TRANSPORT GASIFIER/SULFATOR/HGCU/W501G GT
(O2-BLOWN, 85% Sulfur Capture)

FIGURE 3B

TRANSPORT GASIFIER / SULFATOR / HGCU / W501G GT / 3 PRES LEVEL STEAM CYCLE
(O2-BLOWN, 85% GASIFIER SULFUR CAPTURE)

SUMMARY:

POWER	MWe	EFFICIENCY	%
GAS TURBINE	272.6	HHV	47.1
STEAM TURBINE	142.4	LHV	48.8
MISCELLANEOUS	19.4		
AUXILIARY	11.9		
NET POWER	383.7		

STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13	14
FLOW (LB/HR)	238434	223074	29943	211920	6477	147764	147764	5006006	74122	74122	4393306	612682	612682	42158
TEMPERATURE (F)	59	200	200	630	1290.6	60	180.3	1659.9	59	159.9	1650	1659.9	1004	1004
PRESSURE (PSIA)	14.7	14.7	14.7	400	356	92	400	395.2	14.7	25	400	395.2	385.2	367
H (MM BTU/HR)	-0.8	119	-155.6	-1173.8	-1.8	-0.7	3	-5825.7	-3.1	-1.3	-4379.1	-1465.9	-1653.3	-53.4

STREAM	15	16	17	18	19	20	21	22	23	24	25	26	28	29
FLOW (LB/HR)	570524	2108	579252	111	578587	578143	17498	17498	17498	5250	10849	1400	567183	4030258
TEMPERATURE (F)	1004	995.5	995.5	1079.5	992.5	1084.3	1079.5	487.1	549.4	549.4	549.4	549.4	1078.3	2582.2
PRESSURE (PSIA)	367	362	362	337	357	347	337	375	750	750	750	750	337	268.5
H (MM BTU/HR)	-1599.9	-2.7	-1630.6	-0.1	-1630.6	-1640.7	-49.7	-54.3	-53.9	-16.2	-33.4	-4.3	-1611.3	-1144.9

STREAM	30	31	32	33	34	35	37	38	39	40	41	42	43	44
FLOW (LB/HR)	4557367	4320000	527109	527109	316809	3462603	310772	309448	620220	620220	44377	70424	54552	4627791
TEMPERATURE (F)	1136.6	59	812.7	600	812.7	812.7	59	203.6	512.8	190	1000	1200	300	1137.6
PRESSURE (PSIA)	15.2	14.6	282.2	276.6	282.2	282.2	14.7	280	280	277	335	16	14.7	15.2
H (MM BTU/HR)	-2849.8	-180.3	76.8	48	46.1	504.2	-13	5.1	50.3	0.5	-56.2	-12	-142.4	-2861.8

STREAM	45	46	47	48	49	77	78	B1	B2	B3	B4	W1	W2
FLOW (LB/HR)	548866	649300	72144	71705	1299592	70000	70000	6037	6037	6037	6037	469758	3551
TEMPERATURE (F)	461.5	1030	1030	1290.6	1084.3	607.3	1050	812.1	120	171.3	166.8	61.5	40
PRESSURE (PSIA)	15.2	347	347	357	352	350	342	280.2	275.2	371	362	91	58.5
H (MM BTU/HR)	-442.1	-2233.1	-248.1	-249.4	-4113.2	-388.6	-372	0.9	-0.2	-0.1	-0.1	-5	-0.1

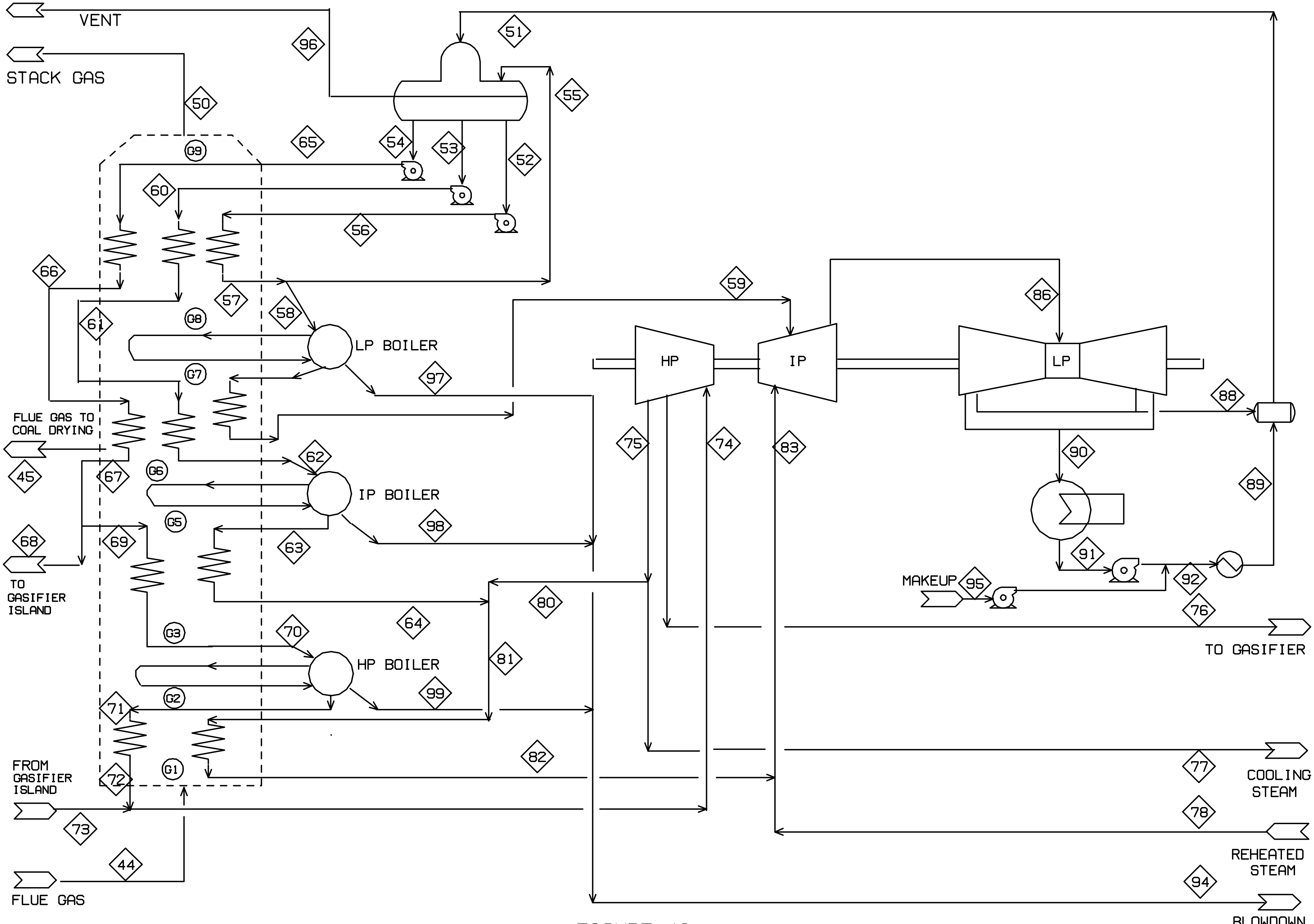

FIGURE 4A

TRANSPORT GASIFIER IGCC - STEAM CYCLE

FIGURE 4B

TRANSPORT GASIFIER / SULFATOR / HGCU / W501G GT / 3 PRES LEVEL STEAM CYCLE
(O2-BLOWN, 85% GASIFIER SULFUR CAPTURE)

STEAM CYCLE / HRSG STREAMS

STREAM	44	45	50	51	52	53	54	55	56	57	58	59	60	61
FLOW (LB/HR)	4627791	548866	4078925	923678	254912	172898	733826	243795	254912	254912	11117	11006	172898	172898
TEMPERATURE (F)	1137.6	461.5	259.7	205	217.3	217.3	217.3	286	217.4	286	286	420	218.2	286
PRESSURE (PSIA)	15.2	15.2	15	17	16.3	16.3	16.3	76.3	80.3	76.3	76.3	68.9	410.6	390
H (MM BTU/HR)	-2861.8	-442.1	-3499.7	-6182.2	-1703	-1155.1	-4902.5	-1611.7	-1702.9	-1685.2	-73.5	-61.9	-1154.8	-1142.9

STREAM	62	63	64	65	66	67	68	69	70	71	72	73	74	75
FLOW (LB/HR)	172898	171169	171169	733826	733826	733826	224421	509405	509405	504311	504311	224421	728732	516812
TEMPERATURE (F)	420	432.3	620	221.1	286	420	420	420	620	629.3	1050	1050	1049.3	607.3
PRESSURE (PSIA)	370.5	352	350	2345.6	2228.3	2116.9	2116.9	2116.9	2011.1	1910.5	1815	1815	1800	350
H (MM BTU/HR)	-1118.6	-969.2	-948.9	-4895.9	-4848.2	-4746.2	-1451.5	-3294.7	-3169.2	-2885.5	-2700.8	-1201.9	-3902.7	-2868.8

STREAM	76	77	78	80	81	82	83	86	88	89	90	91	92	94
FLOW (LB/HR)	211920	70000	70000	446811	617980	617980	687980	698985	53292	870385	645694	645694	870385	6934
TEMPERATURE (F)	638.3	607.3	1050	607.3	610.8	1050	1050	484.3	380.1	140.4	88.8	87.9	80.7	213
PRESSURE (PSIA)	400	350	342	350	350	342	342	35	20	19	0.7	0.7	20	15
H (MM BTU/HR)	-1173.4	-388.6	-372	-2480.3	-3429.2	-3284.4	-3656.4	-3906.4	-300.4	-5881.8	-3771.8	-4397.2	-5933.6	-43.5

STREAM	95	96	97	98	99	G1	G2	G3	G5	G6	G7	G8	G9
FLOW (LB/HR)	224692	5837	111	1729	5094	4627791	4627791	4627791	4627791	4627791	4078925	4078925	4078925
TEMPERATURE (F)	60	217.3	305.3	432.3	629.3	1137.6	887.7	690.4	573.8	461.5	342.5	332.9	259.7
PRESSURE (PSIA)	14.7	16.3	72.5	352	1910.5	15.2	15.2	15.2	15.2	15.2	15.2	15.2	15
H (MM BTU/HR)	-1536.4	-33.4	-0.7	-11.2	-31.6	-2861.8	-3191.2	-3443.3	-3589.1	-3727.2	-3412.2	-3422.4	-3499.7

1. Process Descriptions

Two IGCC Base Cases have been developed based on the Transport gasification process. The cases differ primarily in the oxidant used for the gasification section. Both cases use a raw gas cooler (which is integrated with the gasifier and other heat exchangers) to generate high pressure superheated steam and a ceramic filter to remove particulates, which are recycled to the gasifier. The syngas leaves the gas cooler at 1004°F for both Cases. The fuel gas enters a chloride guard bed that is followed by a hot gas cleanup unit (HGCU) using a transport absorber/regenerator process. The sulfur dioxide rich waste stream from the HGCU is sent to a sulfator. Power is recovered for both cases using a modified W501G gas turbine and a three-pressure level reheat steam cycle.

The composition for the as-received Illinois #6 Coal used in the process is listed below. This coal is dried to approximately 5 % moisture in the coal prep section before being fed to the gasifier.

Proximate Analysis:	**(Wt. %)**	**(Wt. %, dry)**
Moisture	11.12	
Ash	9.70	10.91
Volatiles	34.99	39.37
Fixed Carbon	44.19	49.72
	100	100
HHV (Btu/lb)	11,666	13,126

Ultimate Analysis:	**(Wt. %)**	**(Wt. %, dry)**
Moisture	11.12	
Carbon	63.75	71.72
Hydrogen	4.50	5.06
Nitrogen	1.25	1.41
Chlorine	0.29	0.33
Sulfur	2.51	2.82
Ash	9.70	10.91
Oxygen	6.88	7.75
	100	100

The composition for the sorbent used for sulfur capture in the gasifier is:

	(Wt. %)
CaCO3	97.45
MgCO3	1.58
Inert	0.97
	100.00

Additional features for the both cases are given in following sections. In Table 2, the processes used are compared.

Table 2 : Transport IGCC Base Cases Process Section Comparison

PROCESS SECTION	CASE 1	CASE 2
Gasifier Exit Temp / Press Oxidant	1658 °F/ 395 psia Air	1656°F/ 395 psia Oxygen
Solid Waste **Particulates**	Sulfator Cyclones, Ceramic Filters	same as Case 1
Chloride/NH3 Removal	Chloride Guard Bed	same as Case 1
Sulfur Removal	HGCU - Transport Desulfurization, Sulfator	same as Case 1
Clean Fuel Gas / Gas Addition	Steam	None
Gas Turbine - Power (MWe): - PR / TIT (°F):	Modified W501G 272 (target) 19.37 / 2583	same as Case 1
Steam Cycle - Turb Press: HP/IP/LP - Superheat/Reheat - Exhaust LP Turb - HRSG Stack Temp	3 Pressure Level/Reheat 1800 / 342 / 35 (psia) 1050°F/ 1050°F 0.67 psia 260 °F	same as Case 1

1.1 Transport Gasifier

The Transport Gasifier is a circulating-bed reactor concept, which uses finely pulverized coal and limestone that is proposed by M.W. Kellogg. The gasifier is currently in a development stage, which hopefully will lead to a commercial scale design. It is expected that the small particle size of the coal and limestone will result in a high level of sulfur capture. This may reduce or even eliminate the need for a hot gas cleanup section. Additionally, the small particle size will increase the throughput compared to a KRW gasifier, thereby potentially reducing the required number of gasifier trains (or the gasifier size) and the economic cost.

The Transport Gasifier is conceptually envisioned as consisting of a mixing section , a riser section and a solids recirculation section. The mixing section has a combustion zone and a coal devolatilization zone. The combustion zone is fed with recirculating solids (char, ash, sorbent), oxidant and steam. Sufficient char is burned to provide the heat necessary for the devolatilization of fresh coal and later gasification reactions. The fresh coal and sorbent are injected above the combustion zone and are rapidly heated by the circulating solids and combustion gases in a devolatilization zone. The resulting gas and entrained solids enter the riser section where additional residence time allows the char gasification, methane/steam reforming, water gas shift and sulfur capture reactions to occur. Following the riser section, a solids recirculation section, which includes primary and secondary cyclones, separates the solids into a standpipe system connected with the mixing section. The exiting fuel stream from the secondary cyclone is sent through a raw gas cooler where it is cooled to 1004°F. The heat recovered in the cooler is used in the generation and superheating of high-pressure steam for the steam cycle.

Case 2 differs from Case 1 in that oxygen is used as the oxidant. Because of the much lower flow rate of oxygen compared to air, additional steam is added to maintain the bed circulation and to control the temperature. The steam flowrate used for Case 2 was based on obtaining a maximum combustion zone temperature of approximately 2000°F. Figures 1 and 3 illustrate the gasification section and major process streams relationship to other process sections. In Table 3, gasifier conditions are listed for both of the Transport IGCC cases.

Table 3. Transport IGCC Base Cases - Gasifier Conditions

	CASE 1 Air-Blown	CASE 2 Oxygen-Blown
Coal Flowrate (tons/day)		
- to Prep Plant	2927	2861
- to Gasifier	2739	2677
Coal Moisture (wt. %)		
- to Prep Plant	11.12	11.12
- to Gasifier	5.00	5.00
Gasifier Conditions		
- Exit Pressure (psia)	395	395
- Exit Temp (°F)	1658	1660
- Max. Combustion Temp (°F)	1873	1994
- Ave Riser Temp (°F)	1710	1723
- Carbon Conversion (%)	96.8	98.9
- Sulfur Capture (%)	85.0	85.0
- Residence Time (sec)	3.00	3.00
Flowrates (lb/hr)		
- Coal Feed	228,222	223,074
- Oxidant	657,280 (Air)	147,764 (95%O2)
- Steam	77,596	211,920
- Sorbent	30,634	29,943
- Solids Recycle	4,781,947	4,393,306
- Unrecycled Fines	48,299	44,377
- Raw Fuel Gas	954,045	578,587
Gasifier Flow Ratio:		
- Oxidant/Coal	2.880	0.662
- Steam/Coal	0.340	0.950
- Sorbent/Coal	0.134	0.134
- Recycled Carbon/ Feed Carbon	3.143	1.358
- Recycled Solids/ Feed Solids	18.47	17.364
Heating Value(from gasifier)		
- LHV (Btu/Scf)	145	199

1.2 Gasifier Oxidant Supply

For both Case 1 and Case 2, the oxidant supply required for the gasifier is integrated with the W501G gas turbine. For Case 1, air is bled from the gas turbine compressor exhaust and sent to an air boost compressor. The boost compressor provides the air both for the gasifier and for the regenerator in the HGCU section.

For Case 2, an advanced high pressure cryogenic oxygen plant that takes advantage of the air (278 psia) extracted from the W501G gas turbine supplies oxygen (95% purity) to the gasifier. This advanced design is available due to recent improvements made to the conventional air separation technology which operates efficiently only to about an air supply pressure of 170 psia. The advanced ASU by operating at a higher pressure results in the oxygen and nitrogen products being available from the cold box at higher pressures than in a conventional ASU. This reduces costs for the further compression of these streams. For operational flexibility, (in startup and turndown), the present case considers that the air is supplied, in equal amounts (50%), from a bleed from the gas turbine compressor exhaust and as air supplied directly using a boost compressor. The nitrogen stream produced by the ASU is vented to the atmosphere or available for plant purge nitrogen. (Since a large amount of steam is used in the gasifier for Case 2, a nitrogen recycle to the gas turbine was not required to obtain the desired gas turbine power production of 272 MWe.) Table 4 lists some of the key parameters.

Table 4. Transport IGCC Base Cases - Oxidant Supply Summary

	Case 1 Air-Blown	**Case 2 Oxygen-Blown**
Air to ASU		
- % from Gas Turbine	N/A	50
- % from Boost Compressor	N/A	50
- Flowrate (lb/hr)	N/A	620,220
Air to Boost Compressor		
- % from the Gas Turbine	100	N/A
- Flowrate (lb/hr)	663,316	N/A
Oxidant Stream to Gasifier		
- Flowrate (lb/hr)	657,280	147,764
- Purity (mole % O_2)	20.7	95.0
- ASU O_2 Exit Press (psia)	N/A	92
- Pressure to Gasifier (psia)	400	400
Power Requirements (MWe)		
- Boost Air Compressor*	3.9	13.5
- O_2 Boost Compressor	N/A	2.8

*(Case 1 uses a single stage for air sent to the gasifier, Case 2 uses a five stage intercooled compressor to supply 50% of the air required for the ASU (50%).)

1.3 Chloride Guard Bed / Fine Particulate Removal

The raw fuel gas exits the gas cooler (at 1004°F) and is sent to a cyclone and a gas filter to remove any particulates. This system cleans the gas, leaving the moisture content unchanged, and sends the stream to a chloride guard bed for hydrogen chloride removal. These guard beds containing commercial grade Nahcolite capture the chloride and any other halogens. The beds will require periodic treatment and operate with several on-line while others are being renewed. The resulting fuel gas stream is sent to the HGCU section for sulfur removal. An additional gas filter is used following the HGCU section to guard against any fine particulates left (or generated in HGCU) in the clean fuel gas sent to the gas turbine. A recycle of a small portion of clean fuel gas from the HGCU section is compressed and used for pressurizing gas filters.

1.4 Transport Desulfurization HGCU

The representation for this section was based on information provided by L. Bissett (FETC). FETC is currently developing an on-site (Morgantown) pilot plant to test this HGCU option for a number of sorbents. Since in-bed sulfur capture (85%) was used in the Transport Gasifier , this section serves as a polishing section for sulfur capture. In the HGCU section, the transport absorber operates at an inlet pressure of 357 psia. A zinc titanate sorbent is used. The reaction occurs as a simple exchange between the ZnO portion of the sorbent and the sulfur. The cleaned fuel gas exits and enters a gas filter to capture any particulates before being sent to the gas turbine combustor. (A small portion of the cleaned filtered fuel gas is recycled and pressurized for use in the gas filter.)

The absorber consists of a riser reaction section, a solids/gas separation vessel, and a solids return dipleg. The riser operates at a high void fraction of approximately 95 percent. The large amount of sorbent recirculation results in only a small change in the sorbent sulfur content through this section. A slip stream of approximately 10 percent of the sorbent stream exiting the separation vessel is sent to a regenerator riser, while the remaining portion is combined with regenerated sorbent and sent back for the next absorber cycle. The regenerator is assumed to remove only a portion of the absorbed sulfur. This removal matches the sulfur that is removed from the raw fuel gas that enters the absorber. Since only a small amount of sulfur reacts, the regenerator exit temperature can be controlled to a value less than 1400°F by adjusting the amounts of air (from GT) used. In both cases, the regenerator waste gas stream is sent to the sulfator for disposal. HGCU conditions are listed in Table 5.

1.5 Sulfator

Any SO_2 in the regeneration waste gases, along with any calcium sulfide in the solids from the gasifier, react in the sulfator to form calcium sulfate. Additionally, any unconverted carbon remaining in these streams is oxidized to CO_2. The solid stream, now in an environmentally acceptable form, is cooled for disposal. The sulfator is operated at low pressure (25 psia) with sufficient excess air (10 - 20% excess) supplied directly using an air blower. Heat recovered in the sulfator section is used to generate steam. The exhaust gas from the sulfator is added to the exhaust gas from the gas turbine before entering the steam cycle.

Table 5. Transport Gasifier IGCC Base Case - HGCU Conditions

Sulfur Balance Information:		
Flowrate (lb/hr)	**CASE 1**	**CASE 2**
Sulfur in Raw Fuel Gas	917.16	896.54
Sulfur in Regenerator Waste	892.98	877.24
Sulfur in Clean Fuel Gas	12.73	9.52
(ASPEN Convergence Error Sulfur %)	1.2	1.09
PPMV of Sulfur in Raw Fuel Gas	717.8	944.0
PPMV of Sulfur in Clean Fuel Gas	9.9	9.9
HGCU Sulfur Capture Eff. (weight %)	97.4	97.8
Mole % SO_2 in Regenerator Waste	14.3	14.0
Regenerator Exit Gas Temp (°F)	1338	1291
Regenerator Air Temp (°F)	640	167

HGCU Solids:	**Case 1**		**Case 2**	
	Flowrate (1000 lbs/hr)	**Utilization***	**Flowrate (1000 lbs/hr)**	**Utilization***
To Absorber Rise	721.00	0.4435	721.00	0.4436
From Absorber Separator	721.44	0.4500	721.44	0.4500
To Regenerator Riser	72.14	0.4500	72.14	0.4500
From Regenerator Separator	71.70	0.3849	71.71	0.3861
Ratio: Solids				
- to Absorber/to Regenerator	9.99		9.99	

* Sorbent utilization = moles of ZnS/total moles of ZnX compounds

1.6 Gas Turbine

Both cases were based on using a modified W501G gas turbine. In Case 1 (air-blown gasifier), 15.2% of the compressor discharges is sent to a boost compressor to provide air for the transport gasifier and the HGCU regenerator. In Case 2 (oxygen-blown gasifier), 7.5% of the compressor discharge is used to furnish 50% of the high pressure inlet air for the air separation plant (ASU) and all of the air for the HGCU regenerator. For both Case 1 and Case 2, an additional bleed, 14% of the compressor discharge air, is chilled to 600°F and used for cooling in the turbine expander. Heat recovered from the air cooler is used in the steam cycle. The remainder of the compressor discharge air is used to combust the clean fuel gas. For Case 1, steam is added to increase the flowrate and the power generated in the turbine expander. The steam flowrate is set by requiring that the gas turbine power generated equals approximately 272 MWe. For Case 2, no additional gas (either steam or nitrogen from the ASU) was required to obtain the desired gas turbine power. This is due to a large amount of steam being added directly to the gasifier which is still in the fuelgas stream. In both Case 1 and Case 2, combustor duct cooling is accomplished using intermediate pressure steam supplied from the steam bottoming cycle. This reheated steam is returned to the steam cycle. The combustor exhaust gases enter the expander (2583°F, 269 psia), where energy is recovered to produce power.

The original turbine design specifications are based on a natural gas fuel rather than a coal derived syngas. The syngas's significantly lower heating value when compared to natural gas requires a higher flow rate to obtain the desired turbine firing temperature. To allow for the higher flow rate, an increase in the first nozzle areas will be required. The original combustor will also be replaced with a modified design to handle the lower BTU syngas. In the cases considered, the syngas composition varies depending on the fuel processing prior to the gas turbine and the amount of steam added. In Table 6, the fuel gas composition for each case is listed. In Table 7, the gas turbine conditions are listed for the both Cases.

Table 6. Transport IGCC Base Cases - Fuel Gas Composition (Mole %)

Transport Gas Cleaning	**CASE 1 Air-Blown**	**CASE 2 Oxygen-Blown**
Mole %:		
N_2	44.84	0.66
Ar	0.53	0.46
H_2	19.35	38.51
CO	21.11	19.26
CO_2	9.03	21.66
H_2O	2.63	16.36
CH_4	2.46	3.02
H_2S/COS	9.5 ppmv	9.9 ppmv
NH3	0.05	0.07
Heating Value (HHV) (Btu/Scf)	155	217

Table 7. Transport IGCC Base Cases - W501G Gas Turbine Conditions

Transport Gas Cleaning	**CASE 1 Air-Blown**	**CASE 2 Oxygen-Blown**
Pressure (psia)		
- to Filter	14.7	*(Same
- Compressor inlet	14.57	as Case 1)
- Compressor outlet	282	*
- Combustor exit	269	*
- Expander exhaust	15.2	*
Pressure Ratio	19.4	*
Flowrates (lb/hr)		
- Compr inlet Air	4,320,000	*
- Fuel Gas	944,560	576,183
- Steam	24,000	N/A
- Bleed Air to Gasifier	657,280	N/A
- Bleed Air to HGCU	6,036	6,037
- Bleed Air to ASU	N/A	310,772
- Air Cooling Bleed	527,109	*
- Air Compr Leakage	13,478	*
- Steam Combustor Duct Cooling	70,000	*
- Expander Exhaust	4,611,765	4,557,367
- Gas to HRSG	4,721,818	4,627,791
Temperature (°F)		
- Inlet Air	59	*
- Compressor outlet	810	*
- Steam	602	N/A
- Fuel Gas	1047	1078
- Combustor exhaust	2613	2611
- Turbine inlet	2584	2582
- Turbine exhaust	1128	1137
Power (MWe)		
- Compressor	-237.8	-237.2
- Expander	514.5	513.7
- Generator Loss	-3.9	-3.9
- Net Gas Turbine	272.8	272.6

1.7 Steam Cycle

The steam cycle used for the Cases is based on a design by D. Turek (ABB Power Plant Laboratories). Pressure drops and steam turbine isentropic efficiencies were based on information from a study by Bolland[1]. The cycle is a three-pressure level reheat process. Major components include a heat recovery steam generator (HRSG), steam turbines (high, intermediate, and low pressure), condenser, steam bleed for gasifier steam, steam bleed for gas turbine cooling, recycle water heater, and deaerator. The major difference between Case 1 and Case 2 is related to the amount of steam bleed for the transport gasifier. Case 1 (air-blown gasifier) uses 0.34 lb Steam/lb Coal, while in Case 2 (oxygen-blown gasifier) steam usage increases to 0.95 lb Steam/lb Coal. This results in the net steam power generation being reduced by 20.2 MWe in Case 2 when compared to Case 1. An additional difference is that steam is sent to the gas turbine combustor (24,000 lb/hr at 350 psia) to increase the mass flowrate in the expander to obtain the target of 272 MWe for the gas turbine power production.

In Figures 2 and 4 the steam cycle and process flows are provided for the cases. The primary heat recovered is from the exhaust gas stream of the gas turbine, the sulfator section, and the syngas cooler for the raw fuel gas exiting the gasifier. Additionally, heat is integrated from the gas turbine cooling air chiller, from recycle gas coolers, and from several gasifier island gas coolers. Steam generation occurs at the three pressure levels of 72.5 psia, 353 psia, and 1911 psia in the HRSG. The cycle includes a parallel superheating/reheating section that raises the temperature to 1050°F for both the high-pressure steam and for the combined intermediate pressure steam and high-pressure turbine exhaust steam. Gasifier steam is provided using a bleed from the HP turbine at a pressure of 400 psia. Steam for the gas turbine combustor duct cooling is extracted from the HP turbine at a pressure of 350 psia. The return steam from the gas turbine combustor is combined with reheat steam and sent to the IP steam turbine. The LP steam turbine discharges at 89°F and 0.67 psia. The steam cycle conditions are summarized in Table 8.

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1 "A Comparative Evaluation of Advanced Combined Cycle Alternatives," Transactions of the ASME, April 1991.

Table 8. Transport IGCC Base Cases - Steam Cycle Conditions

HRSG Stack Gas Temperature: 260 °F
Deaerator Vent: 0.5% of inlet flowrate
LP,IP, and HP drum blowdown: 1.0% of inlet flowrate
Pressure drops: 5% of inlet (except IP superheater - 2 psia and line Drop before HP turbine - 15 psia)
High Pressure Turbine Inlet: 1800 psia / 1050 °F
Intermediate Pressure Turbine Inlet: 342 psia / 1050 °F
Low Pressure Turbine Inlet: 35 psia
Low Pressure Turbine Exhaust: 0.67 psia

Pressure Level	**Steam Conditions**		**HRSG Approach Delta Temp (°F)**	
	Pressure (psia)	**Saturation Temp (°F)**	**CASE 1**	**CASE 2**
Low	72.5	305	30	28
Intermediate	352	432	34	30
High	1911	629	62	61

Power Production (MWe)	**CASE 1 Air-Blown**	**CASE 2 Oxygen-Blown**
Steam Turbines	165.1	144.6
Generator Loss	-2.5	-2.1
Net Steam Turbines	162.6	142.4
Pump	-2.1	-2.2

1.8 Power Production

An auxiliary power consumption is assumed as 3 percent of the total power production by the Gas Turbine and the Steam Turbines minus the power consumed by the miscellaneous pumps, expanders, compressors, and blowers. The power production and the overall process efficiency are listed in Table 9 for the both Transport IGCC cases.

Table 9. Transport IGCC Base Cases - Power Production

	CASE 1 Air-Blown	CASE 2 Oxygen-Blown
Gas Turbine (MWe)	272.8	272.6
Steam Turbine (MWe)	162.6	142.4
Miscellaneous (MWe)	-7.2	-19.4
Auxiliary (MWe)	-12.8	-11.9
Plant Total (MWe)	415.4	383.7
Overall Process Efficiency (HHV, %):	49.8	47.1
Overall Process Efficiency (LHV, %):	51.7	48.8

2. Simulation Development

The major question in the simulation development was the representation of the transport gasification process that is currently in the research stage and is not commercially available. The model used was developed by (S. Venkatesan and M. Jarvis, EG&G, 1995) and is a Fortran code that is incorporated into the ASPEN simulations as a “USER” block. This code was validated using limited data made available from M. W. Kellogg (MWK):

- Pressurized Fluid-Bed Combustion Alternative Advanced Concepts, MWK, Final Report, DOE/MC/25000-2934.
- Gasification & Combustion of Coals and Chars in Kellogg's Transport Reactor Test Unit (TRTU). MWK, Test Report, Vol. 1 Results & Discussion, DOE/DE-FC21-90MC25140.
- Private Communication, MWK, August-October 1994. (See “Transport Reactor Model, Topical Report” by S.Venkatesan & M. Jarvis, 1995, DOE/DE-AC21-90MC26328)

The models for the gas turbine (W501G) and the steam cycle were based on previously developed ASPEN simulations. The remaining process sections (i.e. HGCU, CGCU, Acid Plant) were based on representations available in a number of earlier studies. A search of the ASPEN Archive CMS Library will provide example cases for these process sections.

The ASPEN PLUS (version 10.1) simulation codes are stored in the EG&G’s Process Engineering Team Library.

3. Cost of Electricity Analysis

The cost of electricity for the Transport cases was performed using data from the EG&G Cost Estimating notebook and several contractor reports. The format follows the guidelines set by EPRI TAG. Details of the individual section costs are described below and are based on capacity-factored techniques. The COE spreadsheets are included in Appendix A. All costs are reported in 1st Quarter 1999 dollars.

3.1 Coal Preparation

The coal preparation section includes costs for the receiving, conveying, pulverizing and drying systems. The coal flow rate in Case 1 is 2927 tons per day (Illinois #6 coal), resulting in a section cost of $16.7 million. The coal flow rate for Case 2 is 2861 tons per day, resulting in a cost of $16.4 million.

3.2 Limestone Handling and Receiving

The cost for the limestone handling and receiving section includes hoppers, feeders, conveyors and storage silos. The limestone flow rate is 368 tons/day in Case 1, resulting in a cost of $6.8 million. For Case 2, the limestone flow rate is 358 tons/day, with a resulting cost of $6.7 million.

3.3 Oxygen Plant

The cost for the oxygen plant includes the air separation unit, the air precoolers, the oxygen compressors, and the air compressors. The system uses a high-pressure air separation unit. The oxygen plant for Case 2 produces 1773 tons per day oxygen with a total cost of $35.7 million.

3.4 Transport Gasifier

The cost for the gasification section includes the gasifier and the raw gas cooler. Case 1, with 2 gasifier trains, has a cost of $57.6 million. No similar case was found for the oxygen-blown case and the cost was scaled based on the volumetric flow rate of gas entering the gasifier. Case 2, also with 2 process trains, has a cost of $44.0 million. A process contingency of 20 percent was added to the total plant cost based on the development of the gasifier.

3.5 Gas Conditioning

The gas conditioning section includes cyclone, two gas filters and chloride guard beds. The cost for Case 1 is $20.2 million and is based on 2 process trains. The cost for Case 2 is $16.3 million and is based on 2 process trains. A process contingency of 15% was added to the total plant cost based on the development of the gas conditioning components.

3.6 Desulfurization Section

The cost for the transport desulfurization section was derived from a previous report[2]. This includes costs for sorbent hoppers, transport desulfurizer and cyclones. The amount of sorbent used was based information from the Separations and Gasification Engineering Division of NETL. The cost for the HGCU for Case 1 is $13.6 million and is based on 2 process trains. The cost for the HGCU for Case 2 is $12.1 million and is based on 2 process trains. A process contingency of 15% was added to the total plant cost based on the development of the desulfurization sections.

3.7 Ash Handling/Disposal

The cost for the ash handling and disposal includes conveyors, separators and storage silo. Case 1 has an ash flow of 664 tons/day, resulting in the cost of $4.7 million. Case 2 has an ash flow of 653 tons/day, resulting in the cost of $4.7 million.

3.8 Sulfator

The cost for the sulfator includes hoppers, feeder, sulfator, cyclones and fines combustor. The total cost for Case 1 is $13.7 million. The total cost for Case 2 is $11.4 million. A process contingency of 15% was added to the total plant cost based on the development of the sulfation section.

3.9 Gas Turbine Section

The cost for the W501G gas turbine was derived from the Gas Turbine World 96 Handbook[3]. The cost from the handbook was $185/kW and included all the basic turbine components. A factor of 7% was added for modifications and installation. The gas turbine powers of 272.8 MW_e and 272.6 MW_e, for Case 1 and Case 2, respectively, resulted in an approximate cost of $54 million. A process contingency of 5% was added to the total plant cost based on the development of the modified gas turbines.

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2 "Advanced Technology Repowering," Final Report, Prepared for the U.S. Department of Energy, Morgantown Energy Technology Center, Prepared by Parsons Power Group, Inc. May 1997

3 Gas Turbine World Performance Specifications, annual issue, Pequot Publishing Inc., Fairfield Connecticut.

3.10 HRSG/ Steam Turbine Section

The cost for the steam cycle is based on a three-pressure level steam cycle. Case 1 steam turbine power is 162.6 MW_e, with a combined section cost of $47.2 million. Case 2 steam turbine power is 142.4 MW_e, with a combined section cost of $44.2 million.

3.11 Bulk Plant Items

Bulk plant items include water systems, civil/structural/architectural, piping, control and instrumentation, and electrical systems. These were calculated based on a percentage of the total installed equipment costs. The following percentages were used in this report.

Bulk Plant Item	% of Installed Equipment Cost
Water Systems	5.1
Civil/Structural/Architectural	9.2
Piping	5.1
Control and Instrumentation	2.6
Electrical Systems	8.0
Total	30.0

Table 10, Table 11, and Table 12 show the assumptions used in this COE analysis. The total capital requirement for Case 1 is $484,062,000 or $1165/kW, compared to $496,722,000 or $1295/kW for Case 2. The levelized cost of electricity for Case 1 in constant dollars is 38.1 mills/kWh, compared to 41.9 mills/kWh for Case 2.

Table 10. Capital Cost Assumptions

Engineering Fee	10% of PPC*
Project Contingency	15% of PPC
Construction Period	4 Yrs
Inflation Rate	3%
Discount Rate	11.2%
Prepaid Royalties	0.5% of PPC
Catalyst and Chemical Inventory	30 Dys
Spare Parts	0.5% of TPC**
Land	200 Acres @ $6,500/Acre
Start-Up Costs	
Plant Modifications	2% of TPI***
Operating Costs	30 Dys
Fuel Costs	7.5 Dys
Working Capital	
Coal	60 Dys
By-Product Inventory	30 Dys
O&M Costs	30 Dys

* PPC = Process Plant Cost
** TPC = Total Plant Cost
*** TPI = Total Plant Investment

Table 11. Operating & Maintenance Assumptions

<u>Consumable Material Prices</u>	
Illinois #6 Coal	$29.40/Ton
Raw Water	$0.19 /Ton
MDEA Solvent	$1.45/Lb
Claus Catalyst	$470/Ton
SCOT Activated Alumina	$0.067/Lb
Sorbent	$6,000/Ton
Nahcolite	$275/Ton
Off-Site Ash/Sorbent Disposal Costs	$8.00/Ton
Operating Royalties	1% of Fuel Cost
Operator Labor	$34.00/hour
Number of Shifts for Continuous Operation	4.2
Supervision and Clerical Labor	30% of O&M Labor
Maintenance Costs	2.2% of TPC
Insurance and Local Taxes	2% of TPC
Miscellaneous Operating Costs	10% of O&M Labor
Capacity Factor	85%

Table 12. Investment Factor Economic Assumptions

Annual Inflation Rate			3%
Real Escalation Rate (over inflation)			
O&M 0%			
Coal			-1.1%
Discount Rate			11.2%
Debt	80% of Total	9.0% Cost	7.2% Return
Preferred Stock	0% of Total	0.0% Cost	0% Return
Common Stock	20% of Total	20.0% Cost	<u>4.0% Return</u>
			11.2% Total
Book Life			20 Yrs
Tax Life			20 Yrs
State and Federal Tax Rate			38%
Investment Tax Credit			0%
Number of Years Levelized Cost			10 Yrs

Appendix A

Air-Blown Transport HGCU IGCC CASE 1 — 415 MW POWER PLANT
1st Q 1999 Dollar

Total Plant Investment

AREA NO	PLANT SECTION DESCRIPTION	PROCESS CONT, %	PROCESS CONT, K$	COST, K$ W/O CONT
11	Coal Preparation	0	$0	$16,686
11	Limestone Receiving/Handling	0	$0	$6,809
12	Transport Gasifier (2)	20	$11,522	$57,609
12	Recycle Gas Compression (2)	5	$76	$1,520
12	Air Boost Compressor	0	$0	$6,808
14	Gas Conditioning (2)	15	$3,032	$20,216
14	Transport Desulfurizer (2)	15	$2,035	$13,565
15	Gas Turbine System	5	$2,707	$54,136
15	HRSG/Steam Turbine	0	$0	$47,192
16	Ash Handling System	0	$0	$4,722
16	Sulfator	15	$2,061	$13,738
18	Water Systems	0	$0	$11,452
30	Civil/Structural/Architectural	0	$0	$20,658
40	Piping	0	$0	$11,452
50	Control/ Instrumentation	0	$0	$5,838
60	Electrical	0	$0	$17,963

		Subtotal, Process Plant Cost	$310,364
Engineering Fees			$31,036
Process Contingency (Using cont. listed)			$21,433
Project Contingency,	15	% Proc Plt & Gen Plt Fac	$46,555
		Total Plant Cost (TPC)	$409,388
Plant Construction Period,	4.0	Years (1 or more)	
Construction Interest Rate,	11.2	%	
Adjustment for Interest and Inflation			$51,390
		Total Plant Investment (TPI)	$460,778
Prepaid Royalties			$1,552
Initial Catalyst and Chemical Inventory			$508
Startup Costs			$11,428
Spare Parts			$2,047
Working Capital			$6,449
Land,	200	Acres	$1,300
		Total Capital Requirement (TCR)	$484,062
		$/kW	1165

ANNUAL OPERATING COSTS – CASE 1

Capacity Factor = 85 %

COST ITEM	QUANTITY		UNIT $ PRICE		ANNUAL COST, K$
Coal (Illinois #6)	2,927	T/D	$29.40	/T	$26,698
Consumable Materials					
Water	1,072	T/D	$0.19	/T	$63
HGCU Sorbent	0.13	T/D	$6,000	/T	$238
Limestone	367.6	T/D	$16.25	/T	$1,853
Nahcolite	2.4	T/D	$275	/T	$205
Ash/Sorbent Disposal Costs	664	T/D	$8.00	/T	$1,647
Plant Labor					
Oper Labor (incl benef)	14	Men/shift	$34.00	/Hr.	$4,158
Supervision & Clerical					$2,328
Maintenance Costs	2.2%				$9,007
Royalties					$267
Other Operating Costs					$776
	Total Operating Costs				$45,388
By-Product Credits					
	0.0	T/D	$0.00	/T	$0
____________________	0.0	T/D	$0.00	/T	$0
____________________	0.0	T/D	$0.00	/T	$0
____________________	0.0	T/D	$0.00	/T	$0
	Total By-Product Credits				$0
	Net Operating Costs				$45,388

BASES AND ASSUMPTIONS – CASE 1

A. CAPITAL BASES AND DETAILS

	QUANTITY		UNIT $ PRICE		COST, K$
Initial Cat./Chem. Inventory					
Water	27336	T	$0.19	/T	$5
HGCU Sorbent	56	T	$6,000	/T	$334
Limestone	9374	T	$16.25	/T	$152
Nahcolite	61	T	$275	/T	17
Total Catalyst and Chemical Inventory					$508

Startup costs			
Plant modifications,	2	% TPI	$9,216
Operating costs			$1,567
Fuel			$645
		Total Startup Costs	$11,428
Working capital			
Fuel & Consumables inv	60	days supply	$5,619
By-Product inventory	30	days supply	$0
Direct expenses	30	days	$829
		Total Working Capital	$6,449

B. ECONOMIC ASSUMPTIONS

Project life	20	Years
Book life	20	Years
Tax life	20	Years
Federal and state income tax rate	38.0	%
Tax depreciation method	MACRS	
Investment Tax Credit	0.0	%

Financial structure

Type of Security	% of Total	Current Dollar Cost, %	Current Dollar Ret, %	Constant Dollar Cost, %	Constant Dollar Ret, %
Debt	80	9.0	7.2 5.8	4.6	
Preferred Stock	0	3.0	0.0 0.0	0.0	
Common Stock	20	20.0	4.0	16.5	3.3
Discount rate (cost of capital)			11.2		7.9

Inflation rate, % per year	3.0
Real Escalation rates (over inflation)	
Fuel, % per year	-1.1
Operating & Maintenance, % per year	0.0

C. COST OF ELECTRICITY – CASE 1

The approach to determining the cost of electricity is based upon the methodology described in the Technical Assessment Guide, published by the Electric Power Research Institute. The cost of electricity is stated in terms of 10th year levelized dollars.

	Current $	Constant $
Levelizing Factors		
Capital Carrying Charge, 10th yr	0.179	0.148
Fuel, 10th year	1.091	0.948
Operating & Maintenance, 10th yr	1.151	1.000
Cost of Electricity - Levelized	mills/kWh	mills/kWh
Capital Charges	28.0	23.2
Fuel Costs	9.4	8.2
Consumables	1.5	1.3
Fixed Operating & Maintenance	5.2	4.5
Variable Operating & Maintenance	0.9	0.8
By-product	0.0	0.0
Total Cost of Electricity	45.1	38.1

Oxygen-Blown Transport HGCU IGCC CASE 2 — 384 MW POWER PLANT, 1st Q 1999 Dollar

Total Plant Investment

AREA NO	PLANT SECTION DESCRIPTION	PROCESS CONT, %	PROCESS CONT, K$	COST, K$ W/O CONT
11	Coal Preparation	0	$0	$16,421
11	Limestone Receiving/Handling	0	$0	$6,695
12	Oxygen Plant	0	$0	$35,695
12	Transport Gasifier (2)	20	$8,807	$44,036
12	Recycle Gas Compression (2)	5	$102	$2,037
14	Gas Conditioning (2)	15	$2,441	$16,273
14	Air Boost Compressor	0	$0	$173
14	Transport Desulfurizer (2)	15	$1,821	$12,137
15	Gas Turbine System	5	$2,705	$54,096
15	HRSG/Steam Turbine	0	$0	$44,156
16	Ash Handling System	0	$0	$4,679
16	Sulfator	15	$1,710	$11,400
18	Water Systems	0	$0	$12,638
30	Civil/Structural/Architectural	0	$0	$22,797
40	Piping	0	$0	$12,638
50	Control/ Instrumentation	0	$0	$6,443
60	Electrical	0	$0	$19,824

	Subtotal, Process Plant Cost	$322,139
Engineering Fees		$32,214
Process Contingency (Using cont. listed)		$17,585
Project Contingency, 15	% Proc Plt & Gen Plt Fac	$48,321
	Total Plant Cost (TPC)	$420,259
Plant Construction Period, 4.0	Years (1 or more)	
Construction Interest Rate, 11.2	%	
Adjustment for Interest and Inflation		$52,755
	Total Plant Investment (TPI)	$473,013
Prepaid Royalties		$1,611
Initial Catalyst and Chemical Inventory		$432
Startup Costs		$11,891
Spare Parts		$2,101
Working Capital		$6,373
Land, 200 Acres		$1,300
	Total Capital Requirement (TCR)	$496,722
	$/kW	1295

ANNUAL OPERATING COSTS – CASE 2

Capacity Factor = 85 %

COST ITEM	QUANTITY		UNIT $ PRICE		ANNUAL COST, K$
Coal (Illinois #6)	2,861	T/D	$29.40	/T	$26,094
Consumable Materials					
Water	2,822	T/D	$0.19	/T	$166
HGCU Sorbent	0.10	T/D	$6,000	/T	$180
Limestone	358.2	T/D	$16	/T	$1,806
Nahcolite	2.4	T/D	$275	/T	$205
Ash/Sorbent Disposal Costs	653	T/D	$8.00	/T	$1,620
Plant Labor					
Oper Labor (incl benef)	15	Men/shift	$34.00	/Hr.	$4,455
Supervision & Clerical					$2,446
Maintenance Costs	2.2%				$9,246
Royalties					$261
Other Operating Costs					$815
	Total Operating Costs				$47,294
By-Product Credits					
		T/D	$0.00	/T	$0
____________________	0.0	T/D	$0.00	/T	$0
____________________	0.0	T/D	$0.00	/T	$0
____________________	0.0	T/D	$0.00	/T	$0
	Total By-Product Credits				$0
	Net Operating Costs				$47,294

BASES AND ASSUMPTIONS – CASE 2

A. CAPITAL BASES AND DETAILS

	QUANTITY		UNIT $ PRICE		COST, K$
Initial Cat./Chem. Inventory					
Water	71958	T	$0.19	/T	$14
HGCU Sorbent	42	T	$6,000	/T	$253
Limestone	9134	T	$16	/T	$148
Nahcolite	61	T	$275	/T	$17
Total Catalyst and Chemical Inventory					$432

Startup costs			
Plant modifications,	2	% TPI	$9,460
Operating costs			$1,800
Fuel			$631
		Total Startup Costs	$11,891

Working capital			
Fuel & Consumables inv	60	days supply	$5,502
By-Product inventory	30	days supply	$0
Direct expenses	30	days	$871
		Total Working Capital	$6,373

B. ECONOMIC ASSUMPTIONS

Project life	20	Years
Book life	20	Years
Tax life	20	Years
Federal and state income tax rate	38.0	%
Tax depreciation method	MACRS	
Investment Tax Credit	0.0	%

Financial structure

Type of Security	% of Total	Current Dollar Cost, %	Current Dollar Ret, %	Constant Dollar Cost, %	Constant Dollar Ret, %
Debt	80	9.0	7.2	5.8	4.6
Preferred Stock	0	3.0	0.0	0.0	0.0
Common Stock	20	20.0	4.0	16.5	3.3
Discount rate (cost of capital)			11.2		7.9

Inflation rate, % per year	3.0
Real Escalation rates (over inflation)	
Fuel, % per year	-1.1
Operating & Maintenance, % per year	0.0

C. COST OF ELECTRICITY – CASE 2

The approach to determining the cost of electricity is based upon the methodology described n the Technical Assessment Guide, published by the Electric Power Research Institute. The cost of electricity is stated in terms of 10th year levelized dollars.

	Current $	Constant $
Levelizing Factors		
Capital Carrying Charge, 10th yr	0.179	0.148
Fuel, 10th year	1.091	0.948
Operating & Maintenance, 10th yr	1.151	1.000
Cost of Electricity - Levelized	mills/kWh	mills/kWh
Capital Charges	31.1	25.8
Fuel Costs	10.0	8.7
Consumables	1.6	1.4
Fixed Operating & Maintenance	5.9	5.1
Variable Operating & Maintenance	1.0	0.9
By-product	0.0	0.0
Total Cost of Electricity	49.6	41.9

Appendix B

Modifications made to 1998 IGCC Process System Study

Modifications made to the 1998 IGCC Process System Study

The attached summaries show the results obtained previously for the 1998 IGCC Process System Study and the results obtained based on the changes listed below to the economic analysis and the process simulations.

Economics

The following changes were made to the economic section of the 1998 System Study cases done by EG&G for the Gasification Technologies Product Team.

- The costs were brought to 1st Quarter 1999 dollars.
- The contingencies for several sections were changed to reflect advancements in technology development.
- The operating and maintenance costs were lowered to reflect recent technology improvements and competitive pressure (Annual Energy Outlook 2000).
  The number of operators was lowered.
  The maintenance costs were lowered. This is based on a percentage of the Total Plant cost.
- The cost for the Air Separation Units were updated to reflect recent price quotes from a supply vendor.
- The cost and attrition rate for the sorbent in the Hot Gas Cleanup cases were updated to reflect improvements in the state of the art sorbent development. The Separations and Gasification Engineering Division of NETL provided this information.
- The escalation rate of coal was updated to –1.1% from –0.9% and the price of coal was updated to $29.40/ton from $30.60/ ton per the Annual Energy Outlook 2000 projections.
- Some equipment costs were updated after viewing recent publications and talking to technical experts at NETL.

Process Simulations

The following changes were made to the process simulation section of the 1998 System Study done by EG&G for the Gasification Technologies Product Team.

- For Oxygen-blown gasifiers, the Air Separation Unit (ASU) uses an advanced cryogenic plant designed to take advantage of air being provided from a high pressure gas turbine. This resulted in the nitrogen and oxygen streams from the ASU being sent to boost compressors at higher pressures. This reduces power requirements for these compressors.
- Process Efficiencies for boost compressors and air compressors were based on industry recommended values. This resulted in isentropic stage efficiencies for air and nitrogen compressors of 83% compared with 85-87% being used in the 1998 study. Additionally, the oxygen boost compressor stage efficiency was set at 74% compared to 85% used previously. These modifications increased power requirements and partially eliminated the advantage (for

oxygen-blown systems) of the above change.

- Simulation Codes are all available for use in ASPEN PLUS Version 10.1. (Some of the 1998 cases were in version 9.3).
- The databank for pure component information was changed to “Pure10” which is ASPEN PLUS latest release. Only minor changes in some stream information resulted from this change.
- The ASPEN representation for boost compressors and the air compressor was changed from a series of compressor + intercoolers (ASPEN Blocks “COMPR” and “HEATX”) to a multi-stage intercooled compressor (ASPEN Block “MCOMPR”). The low quality heat available from intercoolers was not used in the steam cycle. This had a minimal effect since most cases have excess low quality heat available.

FY 2000 IGCC Systems Summary Update

* (Contingencies on Hot Gas Cleanup Sections: Gas Conditioning 15/10%, Transport Desulfurizer 15%, Sulfator 15%)

	Texaco	Texaco		Shell		Destec		British Gas/	
	Quench	Radiant +	Convective					Lurgi	
	CGCU	CGCU	HGCU	CGCU	HGCU	CGCU	HGCU	CGCU	HGCU
	CASE 1	CASE 2	CASE 3	CASE 1	CASE 2	CASE 1	CASE 2	CASE 1	CASE 2
Gas Turbine Power (MWe)	272.7	272.4	272.1	272.3	272.5	272.8	272.6	272.6	272.5
Steam Turbine Power (MWe)	152.3	191.7	183.8	188.9	187.6	172.2	171.1	133.4	130.3
Misc. /Aux. Power (MWe)	42.0	51.3	46.3	48.3	47.8	44.4	43.3	31.1	30.7
Total Plant Power (MWe)	382.9	412.8	409.6	412.8	412.4	400.6	400.4	374.9	372.1
Efficiency, HHV (%)	39.7	43.5	46.5	45.7	48.0	45.0	47.6	45.3	49.4
Efficiency, LHV (%)	41.2	45.1	48.3	47.4	49.8	46.7	49.4	47.0	51.3
Total Cap Requirement ($1000)	$500,599	$594,053	$561,229	$566,101	$564,963	$546,993	$538,933	$533,664	$503,640
$/kW	$1,307	$1,439	$1,370	$1,371	$1,370	$1,365	$1,346	$1,423	$1,354
Net Operating Costs ($1000)	$48,411	$49,422	$43,426	$46,969	$42,562	$46,487	$41,888	$46,445	$40,416
COE (mills/kW-H)	42.5	44.3	41.1	42.1	40.7	42.3	40.4	44.5	41.1

	KRW Air-Blown			KRW		Transport		Transport	
	With	/out In-Bed Sulf Captur		Oxygen Blown		Air-Blown		Oxygen-Blown	
	HGCU	CGCU	HGCU	CGCU	HGCU	CGCU	HGCU	CGCU	HGCU
	CASE 1	CASE 2	CASE 3				CASE 1		CASE 2
Gas Turbine Power (MWe)	272.6	272.4	272.8				272.8		272.6
Steam Turbine Power (MWe)	184.8	177.0	174.3				162.6		142.4
Misc. /Aux. Power (MWe)	24.5	25.3	25.5				20.0		31.3
Total Plant Power (MWe)	432.9	424.1	421.6				415.4		383.7
Efficiency, HHV (%)	48.4	44.3	46.3				49.8		47.1
Efficiency, LHV (%)	50.2	45.9	48.0				51.7		48.8
Total Cap Requirement (x1000)	$566,641	$544,961	$550,305				$484,062		$496,722
$/kW	$1,309	$1,285	$1,305				$1,165		$1,295
Net Operating Costs (x1000)	$54,059	$48,032	$43,740				$45,388		$47,294
COE (mills/kW-H)	42.4	40.3	39.5				38.1		41.9

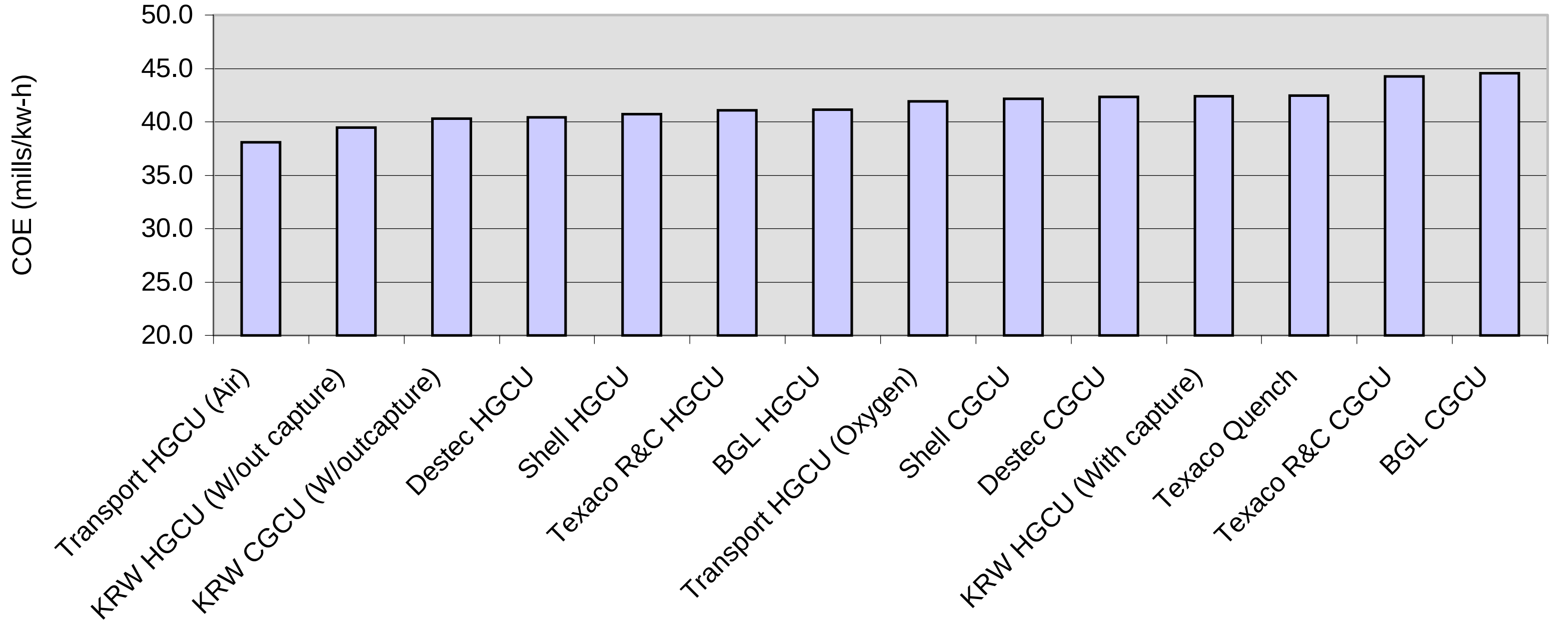
FY 1998 IGCC Systems Summary

	Texaco Quench	Texaco Radiant + Convective		Shell		Destec		British Gas/ Lurgi	
	CGCU CASE 1	CGCU CASE 2	HGCU CASE 3	CGCU CASE 1	HGCU CASE 2	CGCU CASE 1	HGCU CASE 2	CGCU CASE 1	HGCU CASE 2
Gas Turbine Power (MWe)	271.9	272.5	271.2	273.0	271.6	273.0	271.1	272.4	272.1
Steam Turbine Power (MWe)	154.1	192.4	184.9	188.3	189.2	173.5	172.0	131.2	130.7
Misc. /Aux. Power (MWe)	44.4	54.5	49.2	54.3	53.1	48.1	46.3	34.0	33.4
Total Plant Power (MWe)	381.7	410.4	406.9	407.1	407.7	398.5	396.9	369.5	369.3
Efficiency, HHV (%)	39.6	43.4	46.3	45.4	47.5	44.8	47.4	45.4	49.1
Efficiency, LHV (%)	41.1	45.0	48.1	47.0	49.3	46.5	49.1	47.1	50.9
Total Cap Requirement ($1000)	519,625	596,034	593,781	596,811	588,502	551,179	552,513	559,717	528,069
$/KW	1,361	1,452	1,459	1,466	1,443	1,383	1,392	1,515	1,430
Net Operating Costs ($1000)	67,128	69,832	70,836	67,876	69,445	65,711	67,279	65,889	64,710
COE (mills/KW-H)	47.2	48.1	48.8	47.9	48.0	46.2	47.0	50.3	48.5

	KRW Air-Blown With	/out In-Bed Sulf Captur		KRW Oxygen Blown		Transport Air-Blown		Transport Oxygen-Blown	
	HGCU CASE 1	CGCU CASE 2	HGCU CASE 3	CGCU	HGCU	CGCU	HGCU CASE 1	CGCU	HGCU CASE 2
Gas Turbine Power (MWe)	271.8	271.7	272.9				271.4		272.1
Steam Turbine Power (MWe)	181.0	172.7	170.8				160.1		141.9
Misc. /Aux. Power (MWe)	23.8	24.5	24.7				19.5		32.7
Total Plant Power (MWe)	429.0	419.9	419.1				412.0		381.3
Efficiency, HHV (%)	48.4	44.2	46.3				49.9		46.9
Efficiency, LHV (%)	50.2	45.8	48.0				51.7		48.7
Total Cap Requirement ($1000)	607,771	582,832	601,760				520,051		538,369
$/KW	1,417	1,388	1,436				1,262		1,412
Net Operating Costs ($1000)	75,562	68,706	71,722				64,417		67,551
COE (mills/KW-H)	48.3	46.1	48.0				43.6		48.4

COE Summary IGCC Systems Study 2000 Update		COE Summary IGCC Systems Study 1998	
Transport HGCU (Air)	38.1	Transport HGCU (Air)	43.6
KRW HGCU (W/out capture)	39.5	KRW CGCU (W/outcapture)	46.1
KRW CGCU (W/outcapture)	40.3	Destec CGCU	46.2
Destec HGCU	40.4	Destec HGCU	47.0
Shell HGCU	40.7	Texaco Quench	47.2
Texaco R&C HGCU	41.1	Shell CGCU	47.9
BGL HGCU	41.1	KRW HGCU (W/out capture)	48.0
Transport HGCU (Oxygen)	41.9	Shell HGCU	48.0
Shell CGCU	42.1	Texaco R&C CGCU	48.1
Destec CGCU	42.3	KRW HGCU (With capture)	48.3
KRW HGCU (With capture)	42.4	Transport HGCU (Oxygen)	48.4
Texaco Quench	42.5	BGL HGCU	48.5
Texaco R&C CGCU	44.3	Texaco R&C HGCU	48.8
BGL CGCU	44.5	BGL CGCU	50.3

IGCC Base Case COE Comparison



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