Process Engineering Division

British Gas / Lurgi Gasifier
IGCC Base Cases

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PREFACE

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

EXECUTIVE SUMMARY

BRITISH GAS/ LURGI GASIFIER IGCC BASE CASE

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 British Gas/ Lurgi (BGL) gasification process. The objective was to establish a 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 British Gas/ Lurgi published reports. The COE estimates are based on data from the EG&G Cost Estimating Notebook and several contractor reports. These cases can be used as a starting point for the development and analysis of proposed advanced power systems.

The cases developed have the following common process sections:

- Coal Prep - separating, drying, and briquetting of fines.
- BGL Gasification - oxygen-blown, moving-bed, slagging gasifier.
- Air Separation Unit (ASU) - high pressure process integrated with the gas turbine.
- “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 approach used for gas cleanup accounts for the major differences between the two cases. For sulfur removal, Case 1 uses cold gas cleanup (CGCU) and Case 2 uses transport desulfurization hot gas cleanup (HGCU). In Case 1, the raw fuel gas enters a gas/liquor separation and treatment unit where heavy hydrocarbons are removed and then is further cooled, enters a COS hydrolyzer, and is scrubbed (removes remaining particulates, ammonia and chlorides) before entering the CGCU section. In Case 2, the raw fuel gas enters a filter and a chloride guard bed prior to the HGCU section. Sulfur is recovered as elemental sulfur using the Claus process for Case 1 and as sulfuric acid using an acid plant for Case 2. In Case 2, the heavy hydrocarbons remain in the fuel gas and are consumed in the gas turbine combustor.

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 given.

Table 1: British Gas/ Lurgi Gasifier IGCC Base Cases Summary

	CASE 1	CASE 2
Gasifier	BGL	BGL
Sulfur Removal	CGCU	HGCU
Gas Turbine Power (MWe)	272.6	272.5
Steam Turbine Power (MWe)	133.4	130.3
Misc./Ax Power (MWe)	31.1	30.7
Total Plant Power (MWe)	374.9	372.1
Efficiency, HHV (%)	45.3	49.4
Efficiency, LHV (%)	47.1	51.3
Total Capital Requirement, ($1000)	533,664	503,640
$/kW	1,423	1,354
Net Operating Costs ($1000)	46,445	40,416
COE (mills/kWh)	44.5	41.1

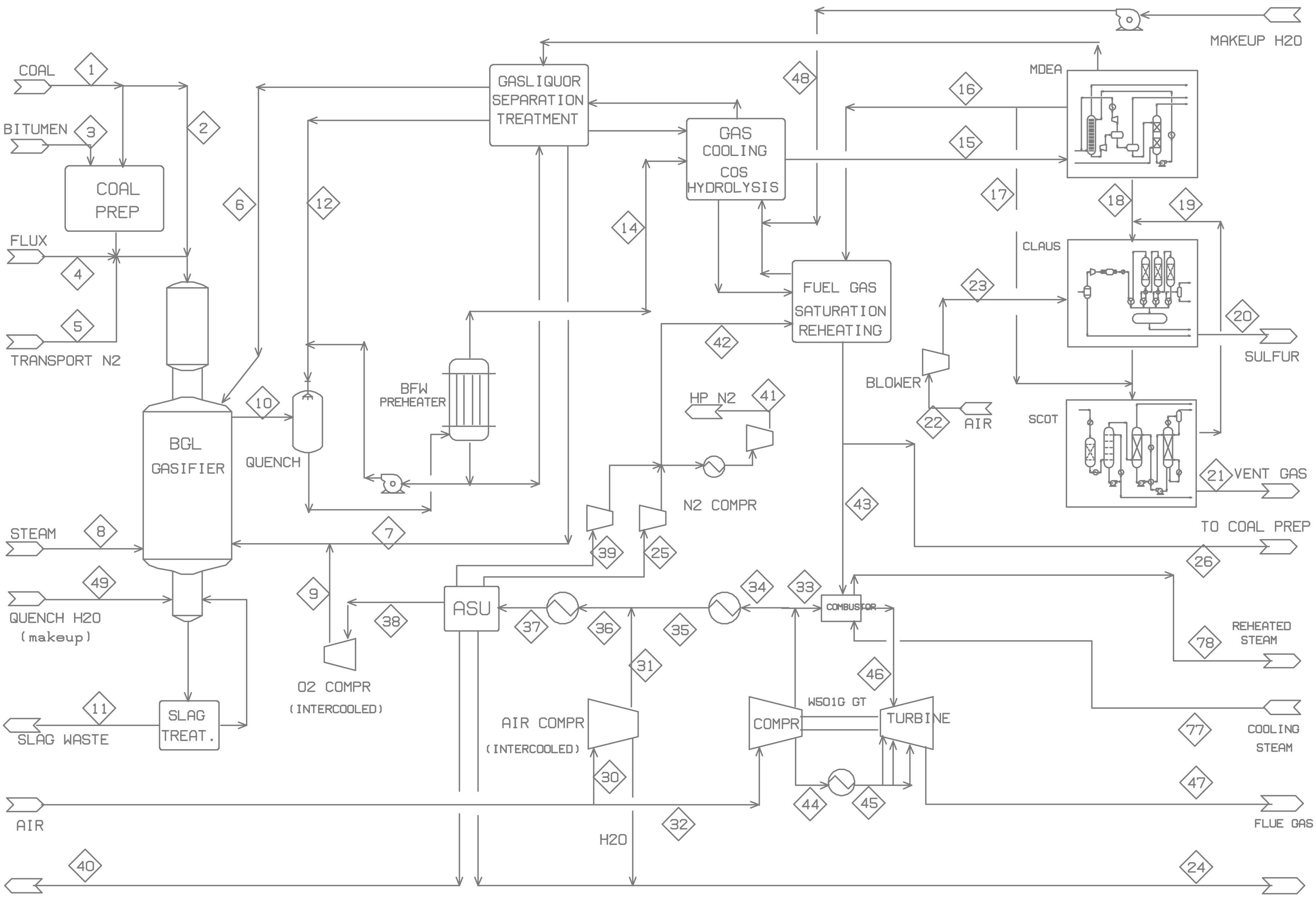

FIGURE 1A
BGL IGCC – CGCU/W501G GT

FIGURE 1B

BGL IGCC - CGCU / W501G GT

SUMMARY:

POWER	MWe	EFFICIENCY:		%
GAS TURBINE	272.6		HHV	45.3
STEAM TURBINE	133.4		LHV	47
MISCELLANEOUS	19.5			
AUXILIARY	11.6			
NET POWER	374.9			

STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
FLOW (LB/HR)	237527	178146	4228	10076	9661	9019	8841	75006	124519	446254	35011	275000	9019	712236	388896
TEMPERATURE (F)	59	59	392	59	110	160	160	684	203.6	986.4	140	176	299	299	100
PRESSURE (PSIA)	14.7	14.7	14.7	14.7	420	395	395	500	464.1	395	14.7	435	392	392	371
H (MM BTU/HR)	-281.3	-211	-1.8	-52.3	0.1	3	3.1	-413.7	3.1	-751.4	-78.5	-1858.5	3.5	-2665.1	-713.5

STREAM	16	17	18	19	20	21	22	23	24	26	30	31	32	33	34
FLOW (LB/HR)	368010	7078	15010	1101	5889	29370	13172	13172	2992	2241	261884	260756	4320000	3517529	261884
TEMPERATURE (F)	116	116	144.2	70	285	70	59	161.2	67.6	550.7	59	204.2	59	813.5	813.5
PRESSURE (PSIA)	376	376	18.5	17.5	14.7	17.5	14.7	25	14.7	335	14.7	280.4	14.7	282.4	282.4
H (MM BTU/HR)	-668.1	-12.8	-37	-3.6	0.3	-73.7	-0.5	-0.2	-20.4	-3.8	-10.9	4.4	-180.3	512.2	38.1

STREAM	35	36	37	38	39	25	40	41	42	43	44	45	46	47	48
FLOW (LB/HR)	261884	522639	522639	124519	126005	25654	244597	14492	137166	573255	527109	527109	4090782	4617892	71442
TEMPERATURE (F)	460	333	179.4	60	62	60	62	112.8	203.8	550.7	813.5	600	2583	1124.1	59.8
PRESSURE (PSIA)	280.4	280.4	280.4	92	91	265	91	425	360	335	282.4	276.8	268.5	15.2	375
H (MM BTU/HR)	14.6	18.9	-0.9	-0.6	-1.4	-0.2	-2.7	0	3.3	-983.5	76.8	47.9	-510.8	-2215.5	-491.8

STREAM	49	77	78
FLOW (LB/HR)	6793	70000	70000
TEMPERATURE (F)	59	606.2	1055.4
PRESSURE (PSIA)	14.7	350	342
H (MM BTU/HR)	-46.8	-388.6	-371.8

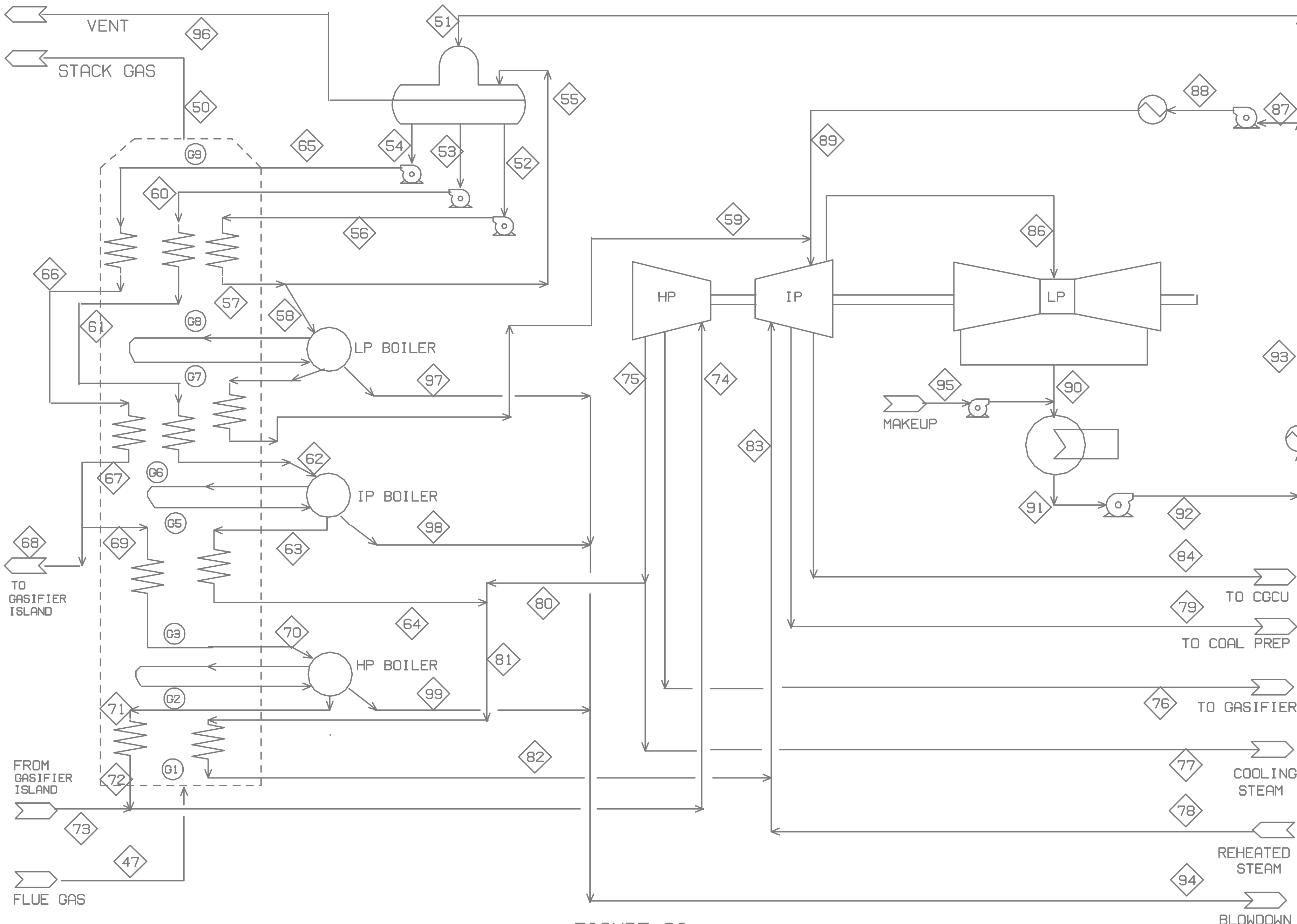

FIGURE 2A
BGL IGCC – STEAM CYCLE

FIGURE 2B

BGL IGCC - CGCU / W501G GT

STEAM CYCLE / HRSG PROCESS STREAMS

STREAM	47	50	51	52	53	54	55	56	57	58	59	60	61	62	63
FLOW (LB/HR)	4617892	4617892	757458	171219	201172	544214	163752	171219	171219	7467	7392	201172	201172	201172	199160
TEMPERATURE (F)	1124.1	292	205	217.3	217.3	217.3	300	217.4	300	300	430	218.1	300	430	432.3
PRESSURE (PSIA)	15.2	14.7	17	16.3	16.3	16.3	76.3	80.3	76.3	76.3	70.5	410.6	390	370.5	352
H (MM BTU/HR)	-2215.5	-3246.8	-5069.7	-1143.9	-1344	-3635.7	-1080.2	-1143.8	-1129.5	-49.3	-41.5	-1343.6	-1326.9	-1299.3	-1127.7

STREAM	64	65	66	67	68	69	70	71	72	73	74	75	76	77	78
FLOW (LB/HR)	199160	544214	544214	544214	62396	481818	481818	476999	476999	62396	539396	464390	75006	70000	70000
TEMPERATURE (F)	620	221.4	300	430	430	430	620	629.3	1050	1050	1049.3	606.2	695.7	606.2	1055.4
PRESSURE (PSIA)	350	2345.6	2228.3	2116.9	2116.9	2116.9	2011.1	1910.5	1815	1815	1800	350	510	350	342
H (MM BTU/HR)	-1104.1	-3630.7	-3587.8	-3513.9	-402.9	-3111.1	-2997.6	-2729.2	-2554.6	-334.2	-2888.7	-2578.1	-413.4	-388.6	-371.8

STREAM	79	80	81	82	83	84	86	87	88	89	90	91	92	93	94
FLOW (LB/HR)	5178	394390	593549	593549	663549	74421	726845	135502	135502	135502	726845	892960	892960	892960	6905
TEMPERATURE (F)	820.3	606.2	610.8	1050	1050.6	598.9	430.5	205	205.1	300	88.8	87.9	87.9	205	213
PRESSURE (PSIA)	150	350	350	342	342	60	35	18	75	65	0.7	0.7	18	18	15
H (MM BTU/HR)	-28.1	-2189.5	-3293.6	-3154.6	-3526.4	-411.9	-4081	-906.9	-906.9	-770.5	-4257.9	-6081.2	-6081.1	-5976.6	-43.4

STREAM	95	96	97	98	99	G1	G2	G3	G5	G6	G7	G8	G9
FLOW (LB/HR)	166116	4606	75	2012	4818	4617892	4617892	4617892	4617892	4617892	4617892	4617892	4617892
TEMPERATURE (F)	60	217.3	305.3	432.3	629.3	1124.1	880.8	690.3	578.5	447	361.2	355.5	292.7
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	15.2
H (MM BTU/HR)	-1135.9	-26.3	-0.5	-13	-29.9	-2215.5	-2529.2	-2767.7	-2904.8	-3063.3	-3165.3	-3172	-3246

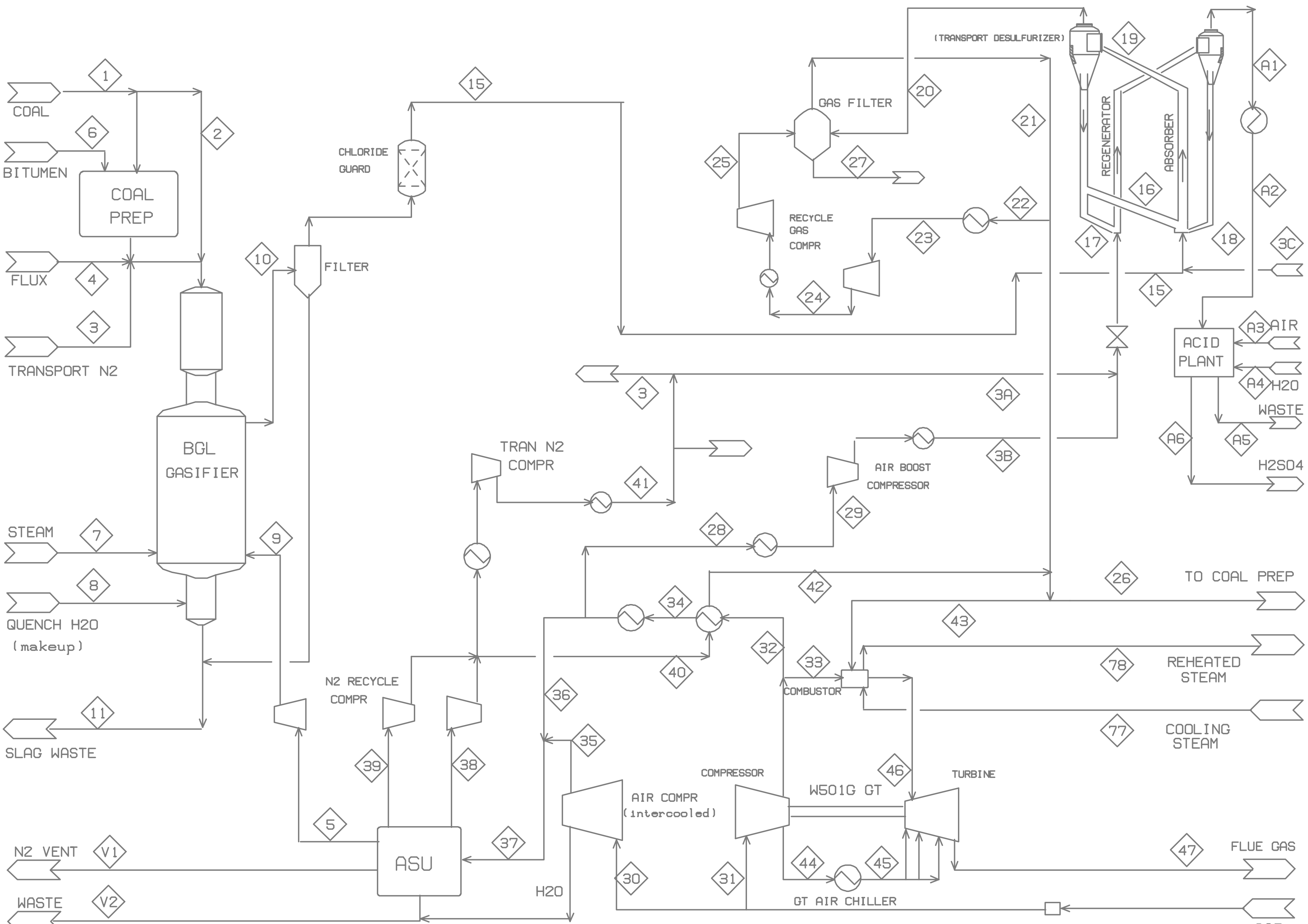

FIGURE 3A
BGL IGCC - (HGCU/W501G)

FIGURE 3B

BGL IGCC - HGCU / W501G GT

SUMMARY:

POWER	MWe	EFFICIENCY:		%
GAS TURBINE	272.5		HHV	49.4
STEAM TURBINE	130.3		LHV	51.3
MISCELLANEOUS	19.2			
AUXILIARY	11.5			
NET POWER	372.1			

STREAM	1	2	3	4	5	6	7	8	9	10	11	15	16	17	18
FLOW (LB/HR)	214558	160919	8821	9174	105509	3850	66256	6185	105509	378604	31878	377984	3744491	416055	413325
TEMPERATURE (F)	59	59	120	59	60	392	700	59	203.4	973.2	140	968.8	1068	1068	1442.1
PRESSURE (PSIA)	14.7	14.7	464	15	92	14.7	500	14.7	464	395	14.7	380	370	370	375
H (MM BTU/HR)	-244.2	-183.2	0.1	-47.6	-0.5	-1.6	-364.9	-42.6	2.6	-666.1	-71.5	-666.2	-12847.6	-1427.5	-1426

STREAM	19	20	21	22	23	24	25	26	27	28	29	3A	3B	3C	V1
FLOW (LB/HR)	4536020	375475	386863	11606	11606	11606	11606	2040	218	35454	35454	16559	35454	218	84786
TEMPERATURE (F)	1070	1070	1065.1	1065.1	750	877.7	905.2	971.2	1065.1	190	120	120	120	100	62
PRESSURE (PSIA)	375	370	360	360	350	501.2	750	294	370	277.2	272.2	464	385	380	91
H (MM BTU/HR)	-14940.3	-667	-687.5	-20.6	-22.2	-21.6	-21.4	-2.2	-0.9	-0.4	-1.1	0.1	-1.2	-1	-0.9

STREAM	V2	30	31	32	33	34	35	36	37	38	39	40	41	42	43
FLOW (LB/HR)	2536	221902	4320000	257356	3522056	257356	220960	221902	442862	21737	229237	221183	29791	221183	594400
TEMPERATURE (F)	59	59	59	812.7	812.7	363	203.7	190	196.8	60	62	192.5	120	712	971.2
PRESSURE (PSIA)	14.7	14.6	14.6	282.2	282.2	280.2	278	277.2	277.2	265	91	300	464	294	294
H (MM BTU/HR)	-17.3	-9.3	-180.3	37.5	512.9	8.1	3.6	-2.5	1.1	-0.1	-2.5	4.7	0.1	34.1	-630.6

STREAM	44	45	46	47	A1	A2	A5	A6	77	78
FLOW (LB/HR)	527109	527109	4116455	4643564	54742	54742	52888	16659	70000	70000
TEMPERATURE (F)	812.7	600	2583	1116.5	1442.1	850	100	100	606.2	1055.4
PRESSURE (PSIA)	282.2	276.6	268.5	15.2	375	365	16	16	350	342
H (MM BTU/HR)	76.8	48	-156.2	-1860.6	-4.5	-12.9	-1	-20.9	-388.6	-371.8

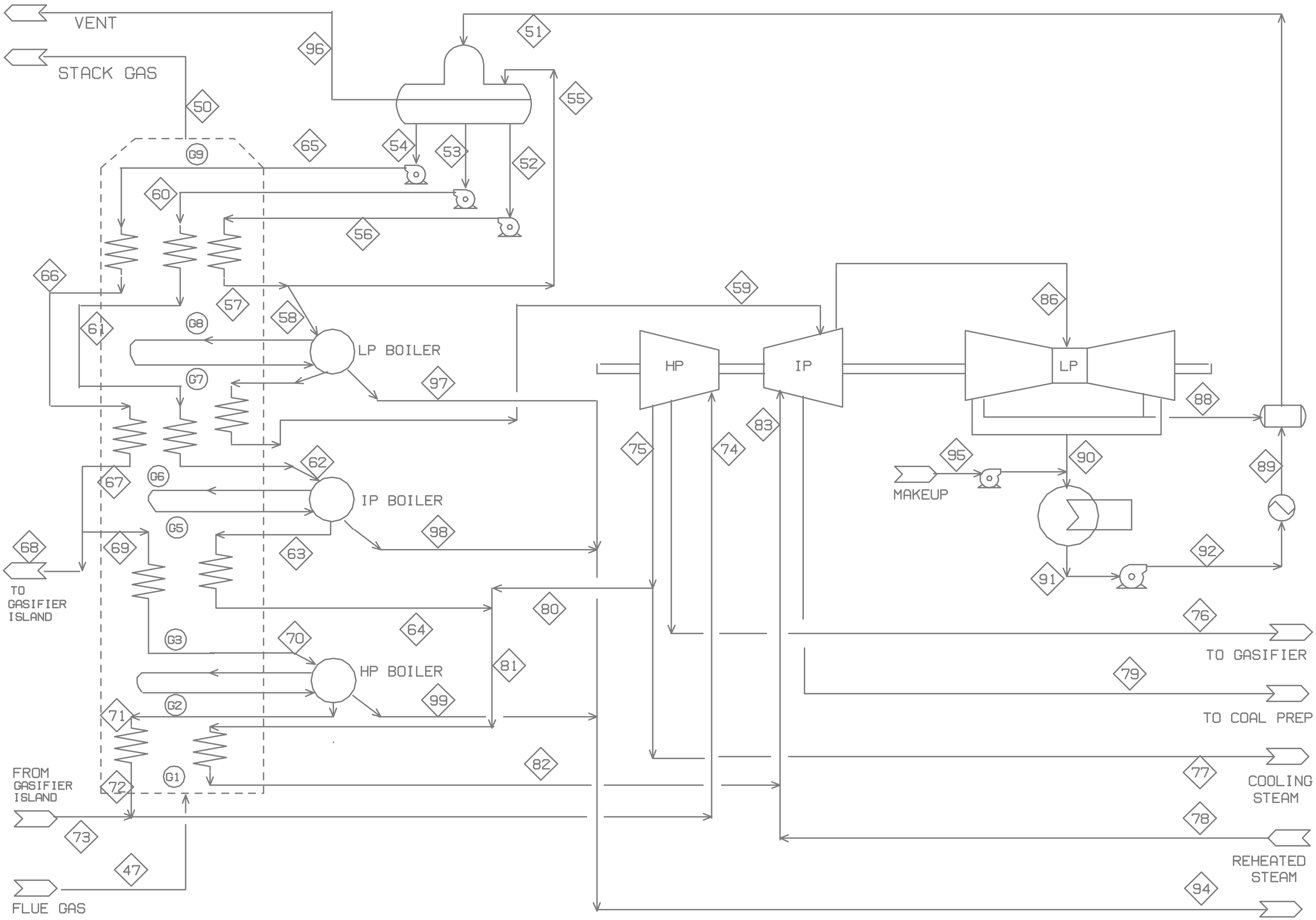

FIGURE 4A
BGL IGCC – STEAM CYCLE

FIGURE 4B

BGL IGCC - HGCU / W501G GT

STEAM CYCLE / HRSG PROCESS STREAMS

STREAM	47	50	51	52	53	54	55	56	57	58	59	60	61	62	63
FLOW (LB/HR)	4643564	4643564	768492	212084	202109	552278	202835	212084	212084	9249	9157	202109	202109	202109	200087
TEMPERATURE (F)	1116.5	292.2	205	217.3	217.3	217.3	286	217.4	286	286	420	218.1	286	420	432.3
PRESSURE (PSIA)	15.2	15	17	16.3	16.3	16.3	76.3	80.3	76.3	76.3	72	410.6	390	370.5	352
H (MM BTU/HR)	-1860.6	-2878.4	-5143.6	-1416.9	-1350.2	-3689.6	-1340.9	-1416.8	-1402.1	-61.1	-51.5	-1349.9	-1336	-1307.5	-1133

STREAM	64	65	66	67	68	69	70	71	72	73	74	75	76	77	78
FLOW (LB/HR)	200087	552278	552278	552278	89574	462704	462704	458076	458076	89574	547651	481395	66256	70000	70000
TEMPERATURE (F)	620	221.4	286	420	420	420	620	629.3	1050	1050	1049.3	606.2	695.7	606.2	1055.4
PRESSURE (PSIA)	350	2345.6	2228.3	2116.9	2116.9	2116.9	2011.1	1910.5	1815	1815	1800	350	510	350	342
H (MM BTU/HR)	-1109.3	-3684.5	-3648.8	-3572	-579.3	-2992.6	-2878.7	-2620.9	-2453.2	-479.7	-2933	-2672.5	-365.2	-388.6	-371.8

STREAM	79	80	81	82	83	86	88	89	90	91	92	94	95	96	97
FLOW (LB/HR)	4714	411395	611482	611482	681482	685924	67880	700613	618044	618044	618044	6741	82569	4857	92
TEMPERATURE (F)	820.2	606.2	610.7	1050	1050.6	482.7	350.7	104.5	88.8	87.9	87.9	213	60	217.3	305.3
PRESSURE (PSIA)	150	350	350	342	342	35	17	17	0.7	0.7	20	15	14.7	16.3	72.5
H (MM BTU/HR)	-25.6	-2283.9	-3393.2	-3249.9	-3621.7	-3833.9	-383.5	-4759.6	-3610.7	-4208.9	-4208.9	-42.4	-564.6	-27.8	-0.6

STREAM	98	99	G1	G2	G3	G5	G6	G7	G8	G9
FLOW (LB/HR)	2021	4627	4643564	4643564	4643564	4643564	4643564	4643564	4643564	4643564
TEMPERATURE (F)	432.3	629.3	1116.5	874.1	690.5	577.9	443.5	354.2	347	292.2
PRESSURE (PSIA)	352	1910.5	15.2	15.2	15.2	15.2	15.2	15.2	15.2	15
H (MM BTU/HR)	-13	-28.7	-1860.6	-2171.6	-2400.6	-2538.3	-2699.8	-2805.6	-2814.1	-2878.4

1. Process Descriptions

Two IGCC Base Cases have been developed based on the BGL gasification process. The BGL process uses an oxygen-blown, moving-bed, slagging gasifier. The Illinois #6 Coal is fed to the gasifier as a mixture of coarse coal (i.e. above 1/4"), fines, and coal briquettes. The oxygen is supplied from a cryogenic air separation plant (ASU) that is integrated with the gas turbine compressor. The steam requirements are furnished from the bottoming steam cycle.

In Case 1, the raw fuel gas produced is cooled to remove the heavy hydrocarbon components (tars, oils, naphtha) which are recirculated to the gasifier. After additional cooling, the fuel gas then enters a cold gas cleanup unit (CGCU) using the MDEA/Claus/Scot process for sulfur removal and recovery. The cleaned fuel gas is reheated, resaturated and combined with recirculated nitrogen from the ASU and sent to the gas turbine section.

In Case 2, the raw fuel gas is sent through a filter, to remove any particulates, and then through a chloride guard bed before being sent to the transport desulfurization unit. The sulfur dioxide generated from the transport regenerator is sent to an acid plant for producing sulfur dioxide. The cleaned hot fuel gas is combined with recirculated nitrogen from the ASU and sent to the gas turbine section.

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 is listed below.

Proximate :	**(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:	**(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 bitumen used as a binder in the briquetting process has the following composition:

Ultimate Analysis:	**(Wt. %)**
Carbon	85.0
Hydrogen	9.7
Nitrogen	1.2
Sulfur	3.4
Oxygen	0.7
	100

HHV (Btu/lb) 17,523

Additional features for the case are given in following sections. In Table 2, overviews of the processes are given.

Table 2 : British Gas/ Lurgi IGCC Base Cases Process Overview

PROCESS SECTION	**CASE 1**	**CASE 2**
Coal Prep **BGL Gasifier** Exit Temp / Press	Sizing, Briquetting 986 °F/ 395 psia	Sizing, Briquetting 973 °F/ 395 psia
Air Separation Plant Inlet Air Pres (psia) O2 Pres (psia) N2 Pres (psia)	 278 92 91 / 265	 278 92 91 / 265
Solid Waste	Slag Quench	Slag Quench
Gas Liquor Separation	Recovers, Recirculates Heavy Hydrocarbons and Removes Particulates.	N/A
Chloride/NH3 Removal	Water Treatment	Chloride Guard Bed
Low Temp Gas Cooling/Heat Recovery	Removes Particulates & Chloride/NH3. Reheating Fuel gas.	N/A
Sulfur Removal	CGCU-MDEA/CLAUS/SCOT (elemental sulfur)	HGCU – Transport Desulfurization, Acid Plant (sulfuric acid)
Clean Fuel Gas / Gas Addition	N2 Recycle from ASU, water resaturating.	N2 Recycle from ASU
Gas Turbine - Power (MWe) - PR / TIT (F)	modified W501G 272 (target) 19.37 / 2583	Modified W501G 272 (target) 19.37 / 2583
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 292 °F	3 Pressure Level/Reheat 1800 / 342 / 35 (psia) 1050°F/ 1050°F 0.67 psia 292 °F

1.1 Coal Prep / BGL Gasifier / Gas Liquor Separation

In the coal prep area, the as-received raw coal is first divided into fractions above and below 1/4" particle size. The coarse coal, along with a portion of the fines, is conveyed to the fuel storage area. The excess fines are sent to the briquetting area, where they are first dried to remove surface moisture. The dryer is fired by the combustion of a small amount of clean fuel gas (bleed from the stream being sent to the gas turbine combustor). After the dryer, the fines are sent to a briquetting press that uses a bitumen binder. The coal briquettes produced are mixed with the coarse coal, fines and flux (required to control slagging in the gasifier) in the fuel storage area.

The BGL gasifier is a counter-current, moving-bed, slagging gasifier operated at pressures of 360 psia or higher. The reactor vessel is water cooled and refractory lined. The coal mixture (coarse coal, fines, briquettes, and flux) is fed into the top of the gasifier via a lockhopper system and reacts while moving downward through the gasifier. The coal's ash/mineral matter is removed from the bottom of the gasifier as molten slag through a slag tap, then quenched in water and removed. Steam and oxygen are injected through tuyere nozzles near the base of the gasifier and react with the coal as the gases move up. This counter-current action results in a wide temperature difference between the top and the bottom of the gasifier. The reactor can be characterized due to this temperature profile as being divided into drying, devolatilization, gasification and combustion zones. Temperatures range from in excess of 3600°F in the combustion zone to an exit product fuel gas temperature of approximately 900°F.

In Case 1, The hot dirty fuel gas leaving the gasifier enters a quench vessel and a BFW preheater designed to lower the temperature to approximately 300°F. Entrained solids and soluble compounds are mixed with the exiting liquid and sent to the gas-liquor separation unit. The soluble hydrocarbon compounds, such as tars, oils, and naphtha are recovered from the aqueous liquor and recycled to the gasifier top and/or reinjected at the tuyeres.

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 the BGL IGCC cases.

Table 3. British Gas/ Lurgi IGCC Base Cases - Gasifier Conditions

	CASE 1	CASE 2
Coal Flowrate (tons/day)		
- Total	2850	2575
- to Dryer/Briquetting	713	644
- Bitumen for Briquetting	51	46
Gasifier Flowrates (lb/hr)		
- Coal Feed (coarse,fines,briquettes)	237,527	214,558
- Flux	10,076	9,174
- Transport Nitrogen	9,661	8,821
- Gasifier Steam	75,006	66,256
- Oxidant	124,519	105,509
- Raw Product Fuel Gas	446,254	378,604
- Tars Recycle	9,019	N/A
- Heavy Hydrocarbons Recycle	8,841	N/A
Gasifier Temperature Profile (°F)		
- Gasifier Exit	986	973
- Devolatilization Zone Exit	1413	1329
- Gasification Zone Exit	1782	1762
- Combustion Zone Exit	3548	3395

1.2 Air Separation Plant (ASU)

For both cases, an advanced high pressure cryogenic oxygen plant that takes advantage of the air (278 psia) extracted from the W501G gas turbine is employed. 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 cases consider 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 GT Compressor bleed air preheats a nitrogen recycle stream (98.9% purity) being sent to the gas turbine to assist in NOX control and to increase the flowrate through the gas turbine expander. The nitrogen recycle is adjusted for each case to yield a net gas turbine power of approximately 272 MWe.. The amount of nitrogen recycle is about 35% for Case 1 and 66% for Case 2. The oxygen stream (95% purity) is supplied to the gasifier. Table 4 lists some of the key parameters for the ASU design.

Table 4. British Gas/ Lurgi IGCC Base Cases - ASU Summary

	Case 1	Case 2
% Air from Gas Turbine	50	50
Air Inlet Press (psia)	278	278
Total Air Flowrate (lb/hr)	522,639	442,862
Oxidant Stream		
- Flowrate (lb/hr)	124,519	105,509
- Purity (mole % O2)	95.0	95.0
- ASU Press (psia)	92	92
- Boost Compr Press (psia)	464	464
Nitrogen Stream		
- Flowrate (lb/hr)	396,256	335,760
- Purity (mole % N2)	98.9	98.9
- ASU Pres (psia)	91 / 265	91 / 265
- Boost Compr Press (psia)	360	300
- % Recycled to GT	35	66
- GT Recycle Temp (°F)	204	712
Power Requirements (MWe)		
- Air Compressor	11.4	9.7
- O2 Boost Compressor	2.9	2.5
- N2 Boost Compressors	2.8	4.3

1.3 Gas Cooling/COS Hydrolysis / Fuel Gas Saturation - Case 1

The raw fuel gas exits the BFW preheater (see Figure 1) in the gasification area and is sent to a gas cooling section. This section cools the raw fuel gas in a series of heat exchangers to a temperature of 100 °F. A liquid waste stream is sent to the Gas Liquor separation process area. Any remaining hydrogen chloride and ammonia is removed in this liquid stream. The cooled raw fuel gas stream is reheated to 285 °F and enters a hydrolysis unit to convert COS to H_2S. This stream is recooled to 100 °F and proceeds to the CGCU section for sulfur removal and recovery. This section's heat recovery is used primarily to preheat water for use in resaturating the clean

fuel gas returning from the CGCU section. Additionally, heat recovered in the CGCU section and from the air coolers preceding the ASU is used for reheating.
The cleaned fuel gas from the CGCU section is combined with a nitrogen recycle stream and resaturated and reheated before being sent to the gas turbine section. This fuel stream is sent to the gas turbine combustor at a temperature of 551 °F with a moisture content of 14.9 % (molar basis). A small bleed of this clean fuel gas is sent to the Coal Prep section for use in coal drying.

1.4 Cold Gas Cleanup Unit (CGCU) - Case 1

The MDEA/Claus/SCOT process is used for cold gas cleanup and sulfur recovery. Refer to Figure 1 for a conceptual idea of the equipment setup for each process. In the MDEA step, the cooled fuel gas from the gas cooling section enters an absorber where it comes into contact with the MDEA solvent. As it moves through the absorber, almost all of the H_2S and a portion of the CO_2 are removed. This clean fuel gas exits the absorber and is sent back to the gas cooling and fuel saturation area described in the preceding report section. The solute-rich MDEA solvent exits the absorber and is heated by the solute-lean solvent from the stripper in a heat exchanger before entering the stripping unit. Acid gases from the top of the stripper are sent to the Claus/SCOT unit for sulfur recovery. The lean MDEA solvent exits the bottom of the stripper and is cooled through several heat exchangers. It is then cleaned in a filtering unit and sent to a storage tank before the next cycle begins.

The Claus process is carried out in two stages. In the first stage, about one-quarter of the gases from the MDEA unit, which exit at 145°F, are mixed with the recycle acid gases from the SCOT unit and are burned in the first furnace. The remaining acid gases are added to the second-stage furnace, where the H_2S and SO_2 react in the presence of a catalyst to form elemental sulfur. The gas is cooled in a waste heat boiler and then sent through a series of reactors where more sulfur is formed. The sulfur is condensed and removed between each reactor. A tail gas stream containing unreacted sulfur, SO_2, H_2S, and COS is sent for further processing in the SCOT unit. This tail gas is heated before entering a reactor where SO_2 converts to H_2S with the aid of a cobalt-molybdate catalyst. The effluent is cooled by waste heat boilers and direct quench before being sent to an absorber column where the H_2S is removed. The H_2S rich stream is sent to the regenerator before being recycled to the absorber. The acid gas from the regenerator is recycled to the Claus step. Further information is provided in Table 5.

Table 5. British Gas/ Lurgi IGCC Base Cases - CGCU Conditions

	Case 1
Sulfur Balance: (lb sulfur/hr)	
- MDEA Feed	5924.2
- Acidgas to Claus	5897.7
- Cleaned Fuel Gas	26.0
- Sulfur Product	5889.0
- SCOT Vent Gas	9.1
Key Conditions	
- PPMV to CGCU	9774
- PPMV Clean Fuel Gas	44.6
- Sulfur Recovery (weight %)	99.6
- Steam Requirements (lb/hr)	74421
- Power Requirements (KWe)	765

1.5 Chloride Guard Bed / Fine Particulate Removal - Case 2

For Case 2, the raw fuel gas exits the particulate removal filter (at 904°F) and is sent to chloride guard bed section 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. A 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.

1.6 Transport Desulfurization HGCU - Case 2

The representation for this section was based on information provided by L. Bissett (NETL). NETL is currently developing an on-site (Morgantown) pilot plant to test this HGCU option for a number of sorbents. In the HGCU section, the transport absorber operates at an inlet pressure of 380 psia. A zinc based 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 at 1003°F 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 of approximately 1400 °F by adjusting the amounts of air (from GT) and nitrogen (from ASU) used. The regenerator waste gas stream is recycled to the sulfuric acid plant for SO_2 removal. HGCU conditions are listed in Table 6.

1.7 Sulfuric Acid Plant - Case 2

In the simulation model, no process details were used to represent the sulfuric acid plant. The only item taken into consideration was the acid plant power consumption rate of 46 watts per lb/hr SO_2 fed to the plant. The sulfuric acid production was based on closing the sulfur balance. However, the following process was used as a basis for the cost analysis.

The regeneration gas from the desulfurization section enters the sulfuric acid plant and passes over a vanadium catalyst stage at temperatures between 800 and 825°F. The temperature is allowed to increase adiabatically as the SO_2 is converted to SO_3. After the reaction is 60 to 70 percent complete, it is stopped. The gas stream is then cooled in a waste heat boiler and passed through subsequent stages of catalyst until the temperature of the gas passing through the last stage is below 800 °F. This process usually requires two to three stages of catalyst. Once cooled, the gas stream is sent to an intermediate absorber tower where some of the SO_3 is removed with 98 percent sulfuric acid. The gases leaving the absorber are reheated and passed over the remaining catalyst stages in a converter. The gases are again cooled and sent to a final absorber tower. Upon exiting the final absorber, the gases are vented to the atmosphere. The conversion of SO_2 to SO_3, and subsequently Sulfuric Acid, using this process is about 99.8 percent.

Table 6. British Gas/ Lurgi Gasifier IGCC Base Cases- HGCU Conditions

Sulfur Balance Information:

	Flowrate (lb/hr)
Sulfur in Raw Fuel Gas	5447.6
Sulfur in Regenerator Waste	5447.2
Sulfur in Clean Fuel Gas	6.0
(ASPEN Convergence Error Sulfur %)	0.103
PPMV of Sulfur in Raw Fuel Gas	9330
PPMV of Sulfur in Clean Fuel Gas	10 (Set in simulation)
HGCU Sulfur Capture Eff. (weight %)	99.89
Mole % SO_2 in Regenerator Waste	9.8
Regenerator Exit Gas Temp (°F)	1442
Regenerator Air Temp (°F)	167

HGCU Solids:	**Flowrate (1000 lb/hr)**	**Sorbent Utilization** *
To Absorber Rise	4157.82	0.443
From Absorber Separator	4160.55	0.450
To Regenerator Riser	416.05	0.450
From Regenerator Separator	413.33	0.381

Ratio: Solids to Absorber/Solids to Regenerator = 9.99

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

1.8 Gas Turbine

The cases were based on using a modified W501G gas turbine that was integrated with the Air Separation Unit (ASU). From the compressor exhaust, a bleed stream is used to supply approximately half of the air supply needed for the ASU. 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. For Case 2, the compressor discharge also supplies air for use in the HGCU regenerator. The remainder of the compressor discharge air is sent directly to the combustor. The fuel gas stream is augmented by the addition of a nitrogen stream supplied from the ASU. This approach is employed to both assist in NOX control and to increase the flowrate which increases the power generated in the turbine expander. The nitrogen recycle flowrate is set by requiring that the gas turbine power generated equals approximately 272 MWe. 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 (2581°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 Table 7, the fuel gas composition for both cases is listed both with and without the nitrogen addition. In Table 8, the gas turbine conditions are listed.

Table 7. British Gas/Lurgi IGCC Base Cases - Fuel Gas Composition (Mole %)

	(No Nitrogen Recycle)		(With Nitrogen Recycle)	
Mole %	**Case 1 CGCU**	**Case 2 HGCU**	**Case 1** CGCU**	**Case 2 HGCU**
O2	-	---	0.1	0.2
N2	2.3	2.1	19.4	31.3
Ar	0.6	0.5	0.5	0.5
H2	29.9	26.7	20.1	18.6
CO	53.8	46.3	36.2	32.3
CO2	5.2	5.9	3.4	4.1
H2O	0.3	10.2	14.9	7.2
CH4	7.5	6.9	5	4.8
H2S	45 PPMV	10 PPMV	26 PPMV	7 PPMV
COS	0.4 PPMV	2 PPBV	0.3 PPMV	1.4 PPBV
C2+	0.5	--	0.3	---
C6+	-	0.4	---	0.3
NH3	-	1.1	---	0.7
Heating Value (HHV) (Btu/Scf)	353	336	238	235

(** Note: For Case 1 the nitrogen is added to the fuelgas, which is then moisturized with water)

Table 8. British Gas/ Lurgi IGCC Base Cases - W501G Gas Turbine Conditions

	CASE 1	**CASE 2**
Pressure (psia)		
- to Filter	14.7	*(Same as Case 1)
- Compressor inlet	14.57	*
- Compressor outlet	282	*
- Combustor exit	269	*
- Expander exhaust	15.2	*
Pressure Ratio	19.4	*
Flowrates (lb/hr)	4,320,000	
- Compressor Inlet Air	3,517,530	*
- Combustor Inlet Air	261,884	3,522,060
- Bleed Air to ASU	N/A	221,902
- Bleed Air to HGCU	527,109	35,454
- Air Cooling Bleed	13,478	*
- Air Compr Leakage	70,000	*
- Steam Combustor Duct Cooling	573,255	*
- Fuel Gas	4,617,890	596,440
- Expander Exhaust Gas to HRSG		4,643,560
Temperature (°F)		
- Inlet Air	59	*
- Compressor outlet	813	*
- Fuel Gas	551	919
- Combustor exhaust	2613	*
- Turbine inlet	2583	*
- Turbine exhaust	1124	1116
Power (MWe)		
- Compressor	-237.2	-237.2
- Expander	513.6	513.6
- Generator Loss	-3.9	-3.9
- Net Gas Turbine	272.6	272.5

1.9 Steam Cycle

The steam cycle used for the two 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 gas turbine cooling, recycle water heater, and deaerator.

In Figure 2 and Figure 4, the steam cycle and process flows are provided for the two cases. The primary heat recovered is from the exhaust gas stream of the gas turbine and from the BGL gasifier section. Additionally, heat is integrated for steam generation from any recoverable gasifier island heat source not required for use in reheating process streams. For example, in Case 2, the heat available from the air compressor bleed air chiller is used in steam generation. However, in Case 1, this heat source is used in reheating the fuel gas and is not available to the steam cycle. Case 1, due to the use of CGCU, has available several sources of medium to low quality heat that are used to both reheat condensate and to generate low pressure steam (see streams 87 - 89, in Figure 2A). Case 2 uses a HGCU for the fuel gas and has available low quality heat to reheat the condensate to only 120°F. Further reheating of the condensate stream to 205°F requires a steam bleed (see stream 88, Figure 4A).

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. Steam is extracted for the gasifier (510 psia) and gas turbine duct cooling (350 psia) from the HP steam turbine. The return steam from the gas turbine duct cooling is combined with reheat steam and sent to the IP steam turbine. An additional steam extraction (150 psia) from the IP turbine is used to provide steam requirements for the CGCU (Case 1 only) and Coal Prep areas. The LP steam turbine discharges at 89°F and 0.67 psia. The steam cycle conditions are summarized in Table 9.

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

Table 9. British Gas/ Lurgi IGCC Base Cases - Steam Cycle Conditions

HRSG Stack Gas Temperature:	292 °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	50	42
Intermediate	352	432	15	11
High	1911	629	61	61

Power Production (MWe)	**CASE 1**	**CASE 2**
Steam Turbines	135.4	132.3
Generator Loss	-2.0	-2.0
Net Steam Turbines	131.4	130.3
Pumps	-1.6	-1.6

1.10 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 10 for the BGL IGCC case.

Table 10. British Gas/ Lurgi IGCC Base Cases - Power Production

	CASE 1 CGCU	**CASE 2 HGCU**
Gas Turbine (MWe)	272.6	272.5
Steam Turbine (MWe)	133.4	130.3
Miscellaneous (MWe)	-19.5	-19.2
Auxiliary (MWe)	-11.6	-11.5
Plant Total (MWe)	374.9	372.1
Overall Process Efficiency (HHV, %)	45.3	49.4
Overall Process Efficiency (LHV, %)	47.1	51.3

2. Simulation Development

The BGL IGCC gasifier island section was developed based on information available in several EPRI reports by BGL. The ASPEN PLUS model for the gasifier was adapted from an earlier DOE sponsored study by Gilbert/Commonwealth. Specifically, the references included:

- DOE Report: (used for ASPEN PLUS gasifier model)
  - ASPEN Model and Economics of a BGC/Lurgi Slagging Gasifier with Hot Gas Desulfurization in a combined-cycle mode, January 1991.
- EPRI Reports:
  - U.S. Bituminous Coal Test Program in the British Gas/Lurgi (BGL) Gasifier, EPRI GS-7091, December 1991.
  - Evaluation of 450-MWe BGL GCC Power Plants Fueled with a Pittsburgh No. 8 Coal, EPRI TR-100376, November 1992.

The models for the gas turbine (W501G), gas cleanup, ASU and the steam cycle were based on previously developed ASPEN Plus Base Case Simulations (FY98 cases). Some sections of the ASPEN PLUS code were based on earlier BGL IGCC ASPEN simulations developed for the DOE. These earlier cases are stored in the ASPEN Archive CMS Library (i.e. CMS BC3-1C and CMS BC3-20C). 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 BGL cases were 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 Handling/ Conveying/ Briquetting

The coal preparation section includes costs for coal hoppers, feeders, conveyors, screening facilities, and briquetting facilities. The coal flow rate in Case 1 is 2850 tons per day (Illinois #6 coal), resulting in a combined section cost of $29.3 million. The coal flow rate in Case 2 is 2575 tons per day (Illinois #6 coal), resulting in a combined section cost of $27.2 million.

3.2 Flux Receiving/Handling

The flux preparation section includes storage and conveying of the flux. The flux flow rate for Case 1 is 121 tons per day, resulting in a section cost of $3.2 million. The flux flow rate for Case 2 is 110 tons per day, resulting in a section cost of $3.0 million.

3.3 Oxygen Plant

The cost for the oxygen plant includes the air separation unit, the air precoolers, the oxygen compressors, the nitrogen compressors and the air compressors. Both systems use a high-pressure air separation unit. The oxygen plant for Case 1 produces 1471 tons per day oxygen with a cost of $32.0 million. The oxygen plant for Case 2 produces 1266 tons per day oxygen with a cost of $29.2 million.

3.4 British Gas/ Lurgi Gasifier

The cost for the gasification section includes the gasifier, the quench unit, the BFW preheater and the slag handling equipment. Both cases are based on two gasification trains. Case 1, with a nominal capacity of 1400 tons per day per train, has a cost of $61.8 million. Case 2, with a nominal capacity of 1300 tons per day per train, has a cost of $57.6 million. A process contingency of 15 percent was added to the total plant cost based on the development of the gasifier.

3.5 Low Temperature Gas Cooling/Fuel Gas Saturation and Gas Liquor Separation and Treatment - CASE 1 only

The cost for the low temperature cooling and gas saturation section includes several heat exchangers, separators, and the saturator column. The cost is $9.8 million.

The gas liquor separation and treatment sections include the tar and oil separators, expansion vessels, product storage tank, strip columns, incinerator, and precipitator. The total cost of the sections is $11.6 million.

3.6 MDEA/ Claus/ SCOT Section - CASE 1 only

The cost of the MDEA acid gas removal system includes the absorber column, the stripping column, heat exchanger and pumps. The cost is $4.7 million.

The cost for the Claus/SCOT sulfur recovery and tail gas treating units is based on 71 tons per day of sulfur entering the unit. The total cost for both units is $13.2 million.

3.7 Gas Conditioning - CASE 2 only

The gas conditioning section includes the gas filters and chloride guard beds. The cost for Case 2 is $11.4 million and is based on two process trains. A process contingency of 15% was added to the total plant cost based on the development of the gas conditioning components.

3.8 Desulfurization Section - CASE 2 only

The cost for the transport desulfurization section was derived from a previous report[2]. This includes costs for sorbent hoppers, transport desulfurizer and cyclones. However, the previous report was for a polishing unit and it is unclear how no sulfur capture in the gasifier will affect the price of the unit or the amount of sorbent needed. The amount of sorbent used was based information from the Separations and Gasification Engineering Division of NETL. The cost for the transport desulfurization section is $9.5 million and is based on two process trains. A process contingency of 15% was added to the total plant cost based on the development of the desulfurization sections.

[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

3.9 Acid Plant Section - CASE 2 only

The cost for the sulfuric acid plant is based on a Monsanto contact process. The unit produces 200 tons per day of sulfuric acid and costs $16.8 million.

3.10 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 Case 1, 272.6 MW_e and Case 2, 272.5 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.

3.11 HRSG/ Steam Turbine Section

The cost for the steam cycle is based on a three-pressure level steam cycle. The steam turbine power for Case 1 is 133.4 MW_e, with a combined section cost of $43 million. The steam turbine power for Case 2 is 130.3 MW_e, with a combined section cost of $42.5 million.

3.12 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 percentages in parenthesis are for the hot-gas cleanup process, which has a lower water requirement, and therefore, a smaller percentage for piping and water systems. The following percentages were used in this report.

Bulk Plant Item	% of Installed Equipment Cost
Water Systems	7.1 (5.1)
Civil/Structural/Architectural	9.2
Piping	7.1 (5.1)
Control and Instrumentation	2.6
Electrical Systems	8.0
Total	34.0 (30.0)

Table 11, Table 12, and Table 13 show the assumptions used in this COE analysis. The total capital requirement for the BGL CGCU Case 1 is $533,664,000 or $1423/kW. The total capital

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

requirement for the BGL HGCU Case 2 is $503,640,000 or $1354/kW. The levelized cost of electricity for the case in constant dollars for Case 1 is 44.5 mills/kWh and for Case 2 is 41.1 mills/kWh.

Table 11. 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 12. Operating & Maintenance Assumptions

Consumable Material Prices	
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 13. 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	4.0% Return
			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

British Gas/ Lurgi CGCU IGCC CASE 1 — 375 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 Receiving/ Handling	0	$0	$14,088
11	Flux Receiving/Handling	0	$0	$3,231
11	Coal Screening/Conveying/Briquetting	0	$0	$15,162
12	Oxygen Plant	0	$0	$32,042
12	BGL Gasifier (2)	15	$9,271	$61,804
14	Gas Liquor Separation	0	$0	$5,284
14	Gas Liquor Treatment	0	$0	$6,355
14	Low Temperature Gas Cooling/Gas Saturation	0	$0	$9,821
14	MDEA	0	$0	$4,665
14	Claus	0	$0	$9,275
14	SCOT	0	$0	$3,922
15	Gas Turbine System	5	$2,705	$54,096
15	HRSG/Steam Turbine	0	$0	$42,950
18	Water Systems	0	$0	$18,651
30	Civil/Structural/Architectural	0	$0	$24,168
40	Piping	0	$0	$18,651
50	Control/ Instrumentation	0	$0	$6,830
60	Electrical	0	$0	$21,016

Subtotal, Process Plant Cost		$352,012
Engineering Fees		$35,201
Process Contingency (Using cont. listed)		$11,975
Project Contingency, 15 % Proc Plt & Gen Plt Fac		$52,802
Total Plant Cost (TPC)		$451,991
Plant Construction Period,	4.0 Years (1 or more)	
Construction Interest Rate,	11.2 %	
Adjustment for Interest and Inflation		$56,738
Total Plant Investment (TPI)		$508,729
Prepaid Royalties		$1,760
Initial Catalyst and Chemical Inventory		$284
Startup Costs		$12,672
Spare Parts		$2,260
Working Capital		$6,659
Land, 200 Acres		$1,300
Total Capital Requirement (TCR)		$533,664
$/kW		1423

ANNUAL OPERATING COSTS – CASE 1

Capacity Factor = 85 %

COST ITEM	QUANTITY		UNIT $ PRICE		ANNUAL COST, K$
Coal (Illinois #6)	2,850	T/D	$29.40	/T	$25,997
Consumable Materials					
Water	2,942	T/D	$0.19	/T	$173
Limestone	121	T/D	$16.00	/T	$600
Bitumen	50	T/D	$130.00	/T	$2,014
MDEA Solvent	403.2	Lb/D	$1.45	/Lb	$181
Claus Catalyst	0.01	T/D	$470	/T	$1
SCOT Activated Alumina	15.9	Lb/D	$0.67	/Lb	$3
SCOT Cobalt Catalyst					$5
SCOT Chemicals					$16
Ash/Sorbent Disposal Costs	413	T/D	$8.00	/T	$1,026
Plant Labor					
Oper Labor (incl benef)	15	Men/shift	$34.00	/Hr.	$4,455
Supervision & Clerical					$2,530
Maintenance Costs	2.2%				$9,944
Royalties					$260
Other Operating Costs					$843
	Total Operating Costs				$48,050
By-Product Credits					
Sulfur	69.0	T/D	$75.00	/T	$1,605
___________________	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				$1,605
	Net Operating Costs				$46,445

BASES AND ASSUMPTIONS – CASE 1

A. CAPITAL BASES AND DETAILS

	QUANTITY		UNIT $ PRICE		COST, K$
Initial Cat./Chem. Inventory					
Water	75024	T	$0.19	/T	$14
Limestone	3083	T/D	$16.00	/T	$49
Bitumen	1273	T/D	$130.00	/T	$166
MDEA Solvent	10282	Lb	$1.45	/Lb	$15
Claus Catalyst	0.3	T	$470	/T	$0
SCOT Activated Alumina	405	Lb	$0.67	/Lb	$0
SCOT Cobalt Catalyst					$16
SCOT Chemicals					$24
Total Catalyst and Chemical Inventory					$284
Startup costs					
Plant modifications,	2	% TPI			$10,175
Operating costs					$1,869
Fuel					$628
	Total Startup Costs				$12,672
Working capital					
Fuel & Consumables inv	60	days supply			$5,603
By-Product inventory	30	days supply			$155
Direct expenses	30	days			$901
	Total Working Capital				$6,659

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

	% of	Current Dollar		Constant Dollar	
Type of Security	Total	Cost, %	Ret, %	Cost, %	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	34.2	28.4
Fuel Costs	10.2	8.8
Consumables	1.7	1.4
Fixed Operating & Maintenance	6.3	5.5
Variable Operating & Maintenance	1.1	1.0
By-product	-0.7	-0.6
Total Cost of Electricity	52.8	44.5

British Gas/ Lurgi HGCU IGCC CASE 2 372 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 Receiving/ Handling	0	$0	$13,122
11	Flux Receiving/Handling	0	$0	$3,026
11	Coal Screening/Conveying/Briquetting	0	$0	$14,122
12	Oxygen Plant	0	$0	$29,241
12	BGL Gasifier (2)	15	$8,636	$57,574
12	Gas Compression (Recycle)	5	$91	$1,815
14	Gas Conditioning (2)	15	$1,707	$11,380
14	Air Boost Compressor	0	$0	$666
14	Transport Desulfurizer (2)	15	$1,422	$9,480
14	Sulfuric Acid Plant	0	$0	$16,759
15	Gas Turbine System	5	$2,704	$54,076
15	HRSG/Steam Turbine	0	$0	$42,520
18	Water Systems	0	$0	$12,943
30	Civil/Structural/Architectural	0	$0	$23,348
40	Piping	0	$0	$12,943
50	Control/ Instrumentation	0	$0	$6,598
60	Electrical	0	$0	$20,302

	Subtotal, Process Plant Cost		$329,915
Engineering Fees			$32,991
Process Contingency (Using cont. listed)			$14,560
Project Contingency, 15 % Proc Plt & Gen Plt Fac			$49,487
	Total Plant Cost (TPC)		$426,953
Plant Construction Period,	4.0 Years (1 or more)		
Construction Interest Rate,	11.2 %		
Adjustment for Interest and Inflation			$53,595
	Total Plant Investment (TPI)		$480,548
Prepaid Royalties			$1,650
Initial Catalyst and Chemical Inventory			$361
Startup Costs			$11,738
Spare Parts			$2,135
Working Capital			$5,908
Land, 200 Acres			$1,300
	Total Capital Requirement (TCR)		$503,640
		$/kW	1354

ANNUAL OPERATING COSTS – CASE 2

Capacity Factor = 85 %

COST ITEM	QUANTITY		UNIT $ PRICE		ANNUAL COST, K$
Coal (Illinois #6)	2,575	T/D	$29.40	/T	$23,487
Consumable Materials					
Water	1,100	T/D	$0.19	/T	$65
HGCU Sorbent	0.05	T/D	$6,000	/T	$97
Nahcolite	3.0	T/D	$275	/T	$256
Limestone	110.0	T/D	$16	/T	$546
Bitumen	46.2	T/D	$130	/T	$1,861
Ash/Sorbent Disposal Costs	382	T/D	$8.00	/T	$949
Plant Labor					
Oper Labor (incl benef)	15	Men/shift	$34.00	/Hr.	$4,455
Supervision & Clerical					$2,464
Maintenance Costs	2.2%				$9,393
Royalties					$235
Other Operating Costs					$821
	Total Operating Costs				$44,630
By-Product Credits					
Sulfuric Acid	199.7	T/D	$68.00	/T	$4,214
___________________	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				$4,214
	Net Operating Costs				$40,416

BASES AND ASSUMPTIONS – CASE 2

A. CAPITAL BASES AND DETAILS

	QUANTITY		UNIT $ PRICE		COST, K$
Initial Cat./Chem. Inventory					
Water	28050	T	$0.19	/T	$5
HGCU Sorbent	23	T/D	$6,000	/T	$136
Nahcolite	77	T/D	$275	/T	$21
Limestone	2805	T/D	$16	/T	$45
Bitumen	1177	T/D	$130	/T	$153
Total Catalyst and Chemical Inventory					$361

Startup costs			
Plant modifications,	2	% TPI	$9,611
Operating costs			$1,560
Fuel			$568
	Total Startup Costs		$11,738

Working capital			
Fuel & Consumables inv	60	days supply	$4,623
By-Product inventory	30	days supply	$407
Direct expenses	30	days	$878
		Total Working Capital	$5,908

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 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	32.5	27.0
Fuel Costs	9.2	8.0
Consumables	1.6	1.4
Fixed Operating & Maintenance	6.1	5.3
Variable Operating & Maintenance	1.1	0.9
By-product	-1.8	-1.5
Total Cost of Electricity	48.8	41.1

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 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)	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 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)	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

June 15, 2000

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

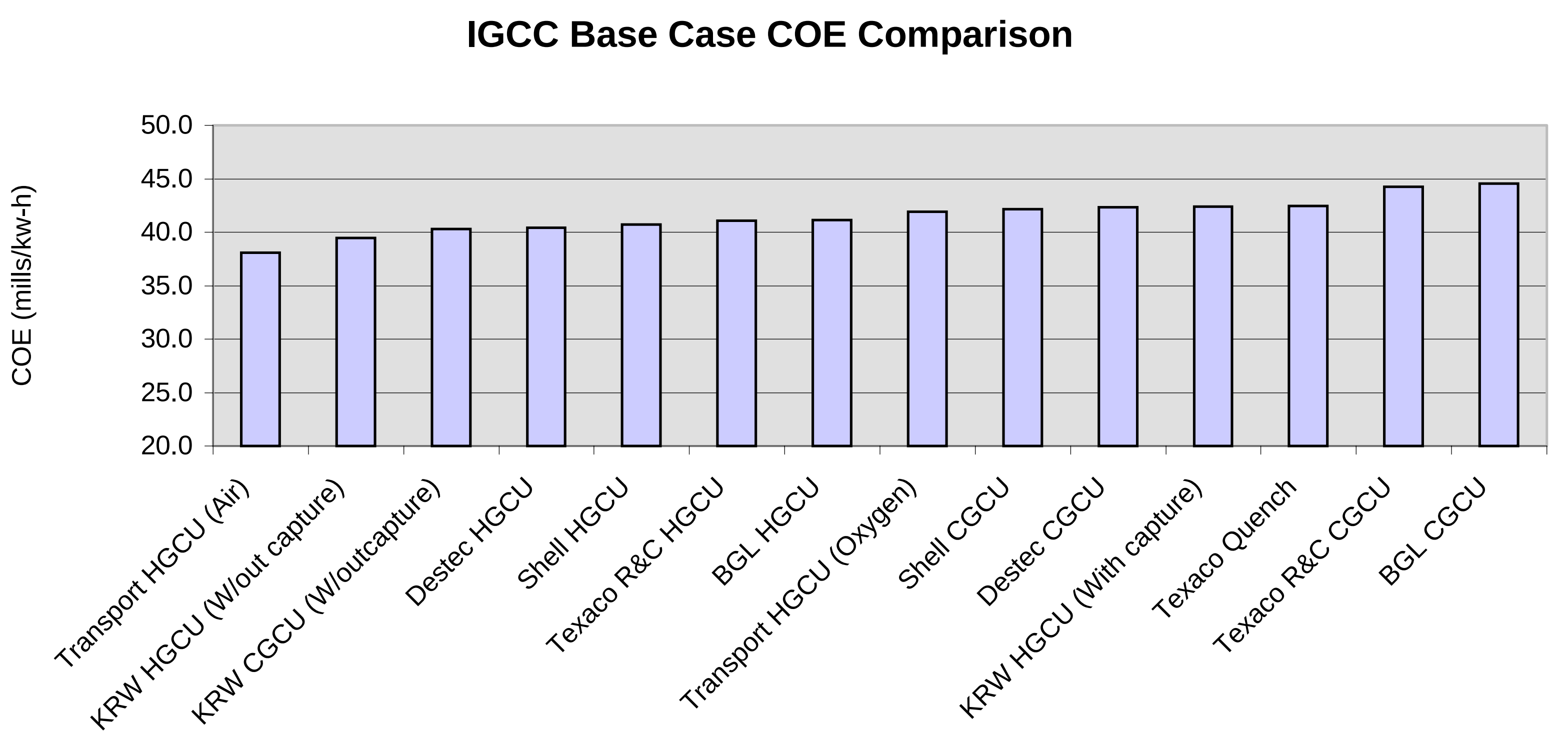
	KRW Air-Blown With	KRW Air-Blown W/out In-Bed Sulf Capture		KRW Oxygen Blown		Transport Air-Blown		Transport Oxygen-Blown	
	HGCU CASE 1	CGCU CASE 2	HGCU CASE 3	CGCU	HGCU	CGCU	HGCU CASE 1	CGCU	HGCU CASE 2
Gas Turbine Power (MWe)	271.8	271.7	272.9				271.4		272.1
Steam Turbine Power (MWe)	181.0	172.7	170.8				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

Transport HGCU (Air)	38.1
KRW HGCU (W/out capture)	39.5
KRW CGCU (W/outcapture)	40.3
Destec HGCU	40.4
Shell HGCU	40.7
Texaco R&C HGCU	41.1
BGL HGCU	41.1
Transport HGCU (Oxygen)	41.9
Shell CGCU	42.1
Destec CGCU	42.3
KRW HGCU (With capture)	42.4
Texaco Quench	42.5
Texaco R&C CGCU	44.3
BGL CGCU	44.5

COE Summary IGCC Systems Study 1998

Transport HGCU (Air)	43.6
KRW CGCU (W/outcapture)	46.1
Destec CGCU	46.2
Destec HGCU	47.0
Texaco Quench	47.2
Shell CGCU	47.9
KRW HGCU (W/out capture)	48.0
Shell HGCU	48.0
Texaco R&C CGCU	48.1
KRW HGCU (With capture)	48.3
Transport HGCU (Oxygen)	48.4
BGL HGCU	48.5
Texaco R&C HGCU	48.8
BGL CGCU	50.3



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