IGCC System Analysis Utilizing Various Condenser Cooling Platforms Including CO_2 Sequestration

Robert E. James III, PhD
US Department of Energy
National Energy Technology Laboratory
Office of Systems and Policy Support
Morgantown, WV

Gilbert V. McGurl, PhD
US Department of Energy
National Energy Technology Laboratory
Office of Systems and Policy Support
Pittsburgh, PA

Abstract

This study details the affect of switching a conventional IGCC plant to various condenser types and adding CO_2 sequestration. Condenser types investigated include a standard once-through water cooled condenser (WCC), a direct air cooled condenser (DACC), and a wet tower condenser (WTC). The sequestration system used incorporates water-gas shift reactors and a two-stage Selexol system for both sulfur and CO_2 removal.

The results show that changing from the once-through WCC to the DACC adversely impacts the system performance, cost, and economics. Addition of the sequestration system to the DACC system further decreases performance and increases costs. The CO2 separated at the plant is sequestered by injection into glacial saline aquifers. Switching out the DACC with a WTC using a non-traditional cooling medium helps the system by regaining some of the lost power, and decreasing costs. The non-traditional cooling medium is saline aquifer water pumped from a deep glacial aquifer. Removal of saline water from the aquifer reduces or eliminates pressure growth that might result from injection of CO2 into the aquifer. This reduces the potential of CO2 leakage from the aquifer.

1.0 Simulation Development

The simulation was run using Aspen Plus version 11.1. The program also allows user input via FORTRAN computer calculation sections to simulate details that would otherwise be missed by Aspen Plus. The system model is set to run with ISO conditions for a Southwest US location. Property sets ranged from Peng-Robinson equation of state, to Steam and Solid tables for the steam turbine and coal handling systems, respectively.

2.0 Process Section Descriptions and Results

The base system used for this study is an IGCC system using a Texaco Gasifier with water quench cooler, generating fuel syngas for use in a Gas Turbine, Case #1 shown in Figure 1. The quench system is a high-pressure water quench section that rapidly reduces the solid/gas mixture to approximately 425 F (605 psia). A gas scrubber and a low temperature, gas cooling/heat recovery section are used to reduce the raw fuel gas stream to 103 F prior to entering a CGCU section for sulfur removal. The CGCU section cleans out sulfur using an MDEA process, producing a low-sulfur fuel gas stream that is sent to the gas turbine.

The Gas Turbine (GT) is a Siemens-Westinghouse 501G machine with turbine inlet temperature of 2583 F. 50% of the air from the GT compressor is sent to the Air Separation Unit (ASU). N_2 from the ASU is used to boost the mass flow through the GT. The GT is bottomed by a 3 pressure steam turbine. The condenser operates at 0.67 psia in a once through water-cooled orientation.

The first revision, Case #2 shown in Figure 2, was effected by changing from a once-through water cooled condenser to a direct air cooled condenser (DACC). Due to location (Southwest US) the ambient

Figure 1. Case #1

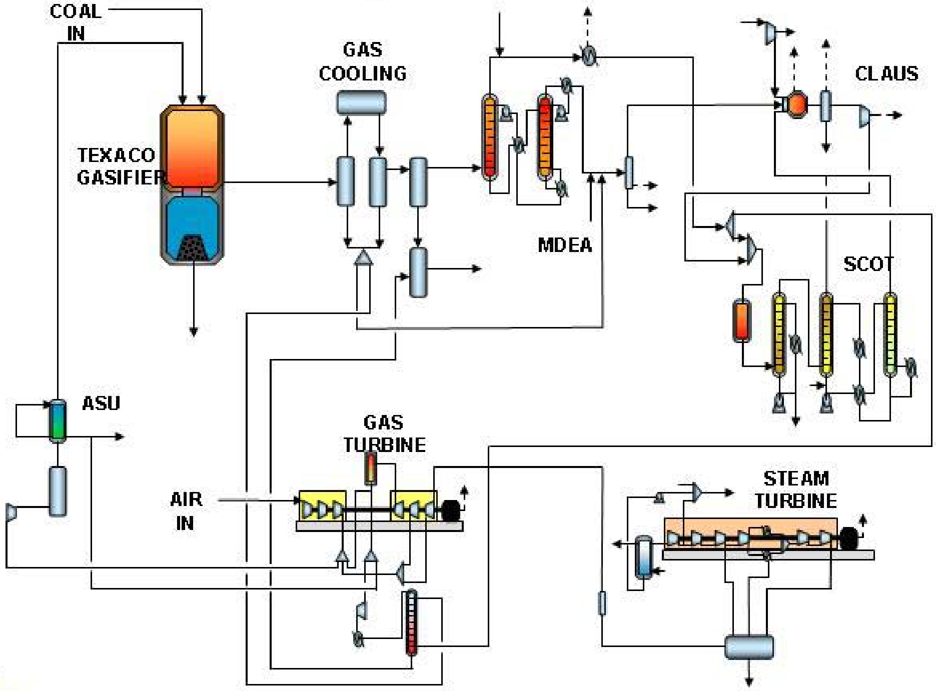
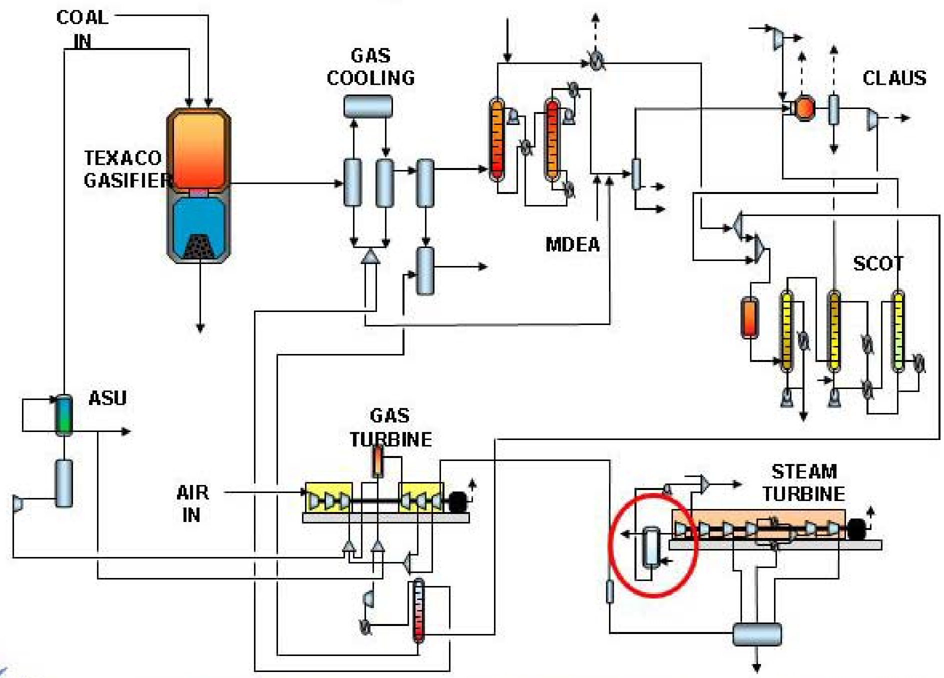


Figure 2. Case #2



temperature forced the condenser to operate at a higher backpressure due to a higher condensate saturation temperature, lowering the power output of the ST and lowering the net power output. The plant footprint also increased due to the large condenser surface area because of the gas-to-liquid interface in the DACC.

In the second revision, Case #3 shown in Figure 3, a pair of water gas shift (WGS) reactors in series were added to convert the syngas to hydrogen and carbon dioxide. In addition, a two-stage Selexol unit was added, taking the place of the MDEA unit and adding a CO_2 removal section. The first stage of the Selexol unit effectively removes sulfur from the syngas stream, sending the sulfur to the Claus and SCOT units. The second stage removes CO_2, which is then stripped from the Selexol yielding a highly concentrated CO_2 stream. The CO_2 is compressed and cooled to liquefy it for sequestration.

In the third revision, Case #4 shown in Figure 4, the DACC is replaced by a Wet Tower Condenser (WTC). The WTC allows lower pressure in the last stage of the ST due to lower condensate saturation temperature than the DACC, hence increasing the net power output. Water used in the cooling tower is from a saline aquifer. The tower has a concentration ratio of 2:1, with blowdown being injected back into the aquifer.

The fuel used for the gasifier systems is a non-specific rank or seam Western Sub-Bituminous coal, assumed to be from the Powder River Basin region. The composition for the as-received Western Sub-Bituminous coal fed to the slurry process is shown in Table 2.1.

Additional features for different cases are given in following sections. Table 2.2 gives details of the process equipment used in all four cases.

2.1 Texaco Gasifier

A generic layout of the Texaco gasifier is shown in Figure 5. Coal is crushed and mixed with water to produce slurry that is 33.5% by weight water (including coal moisture). This slurry is pumped into the gasifier along with oxygen from the ASU. The gasifier operates in a pressurized, downflow, entrained design and gasification takes place rapidly at temperatures in excess of 2300 ^{o}F. The raw fuel gas produced is mainly composed of H_2, CO, CO_2 and H_2O. The coal's sulfur is primarily converted to H_2S and a smaller quantity of COS. This raw fuel gas leaves the gasifier at 2300-2700 ^{o}F, along with molten ash and a small quantity of unburned carbon. No hydrocarbon liquids are generated. This gas/molten solids stream enters the direct quench section.

The Quench design consists of a large water pool that cools the gas and removes solidified ash particles. The cooled raw fuel gas enters a gas scrubbing section to remove additional fine solids before exiting the gasification section to a gas cooling section. Table 2.3 lists gasifier conditions for the cases tested.

2.2 Air Separation Unit (ASU)

For all cases, an advanced high-pressure cryogenic oxygen plant that takes advantage of the air (278 psia) extracted from the W501G gas turbine is employed. 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 flow rate through the gas turbine expander. The nitrogen recycle is adjusted for each case to yield a net gas turbine power of approximately 275 MWe for the base case. The amount of nitrogen recycled is less than 55%. This implies the possibility that two ASU plants could be run in parallel for these cases (a high-pressure oxygen plant with nearly all the nitrogen recycled and a lower pressure oxygen plant with all the nitrogen vented). Table 2.4 lists some of the key parameters for the ASU designs.

Figure 3. Case #3

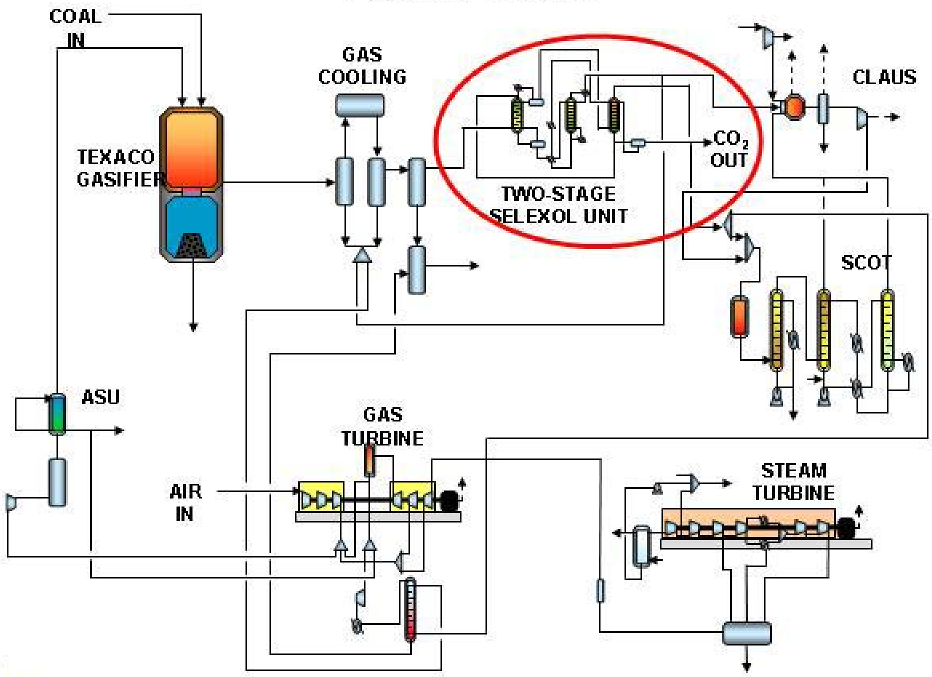


Figure 4. Case #4

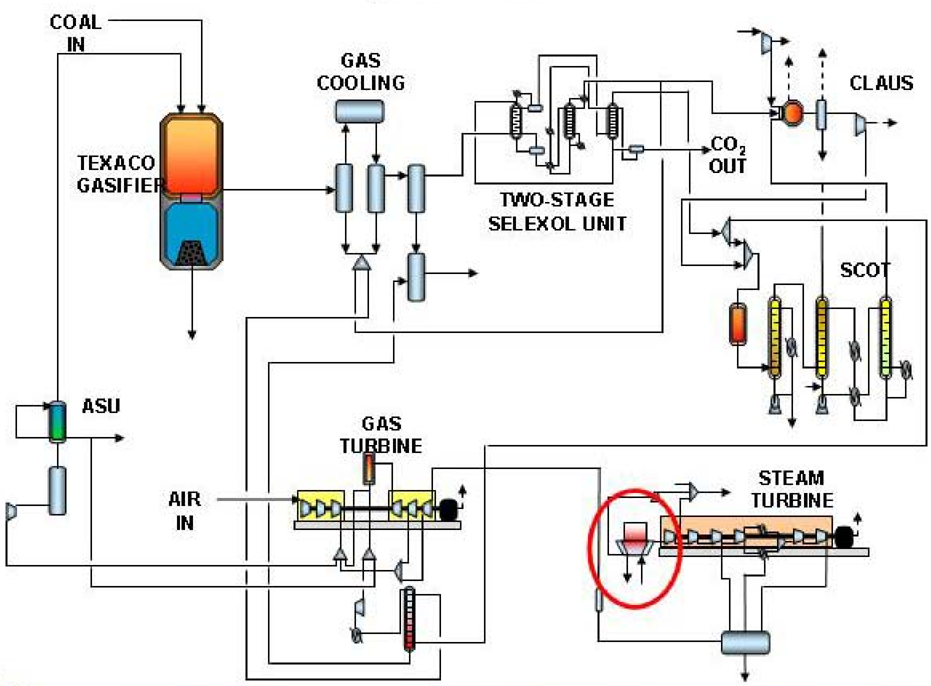


Table 2.1. Coal Spec Analysis

Proximate Analysis	Wt %	Wt % Dry	Ultimate Analysis	Wt %	Wt % Dry	Sulfur Analysis	Wt % Dry
Moisture	30.4		**Moisture**	30.4		**Pyritic**	0.11
Ash	6.4	9.19	**Carbon**	47.85	68.75	**Sulfate**	0.04
Volatiles	31.1	44.68	**Hydrogen**	3.4	4.88	**Organic**	0.33
Fixed Carbon	32.1	46.12	**Nitrogen**	0.62	0.89		
	100	100	**Chlorine**	0.03	0.043		
			Sulfur	0.48	0.69		
			Oxygen	10.82	15.5		
			Ash	6.40	9.195		
HHV (kJ/kg)	8461	12404	**Total**	100	100		
LHV (kJ/kg)	7420	11292					

Table 2.2. Process Section Comparison

Process Section	Case #1	Case #2	Case #3	Case #4
Texaco Gasifier				
Exit Temp/Pres	2500 F/605 psia	2500 F/605 psia	2500 F/605 psia	2500 F/605 psia
Slurry(% Solids)	66.5	66.5	66.5	66.5
Raw Fuel (syngas) Cooling Mode	Quench (425 F)	Quench (425 F)	Quench (425 F)	Quench (425 F)
Air Separation Plant	50% Integration GT	50% Integration GT	50% Integration GT	50% Integration GT
Inlet Air Pres(psia)	277	277	277	277
O2/N2 Pres (psia)	650/336	650/336	650/336	650/336
Solid Waste /Particulates	Slag Treatment, Gas Scrubber	Slag Treatment, Gas Scrubber	Slag Treatment, Gas Scrubber	Slag Treatment, Gas Scrubber
Low Temp Gas Cooling/Heat Recovery	COS Hydrolysis LP & NH_3 Strip Steam BFW Heating	COS Hydrolysis LP & NH_3 Strip Steam BFW Heating	LP & NH_3 Strip Steam BFW Heating	LP & NH_3 Strip Steam BFW Heating
Chloride/NH_3 Removal	Water Condensate Treatment, NH_3 Strip	Water Condensate Treatment, NH_3 Strip	Water Condensate Treatment, NH_3 Strip	Water Condensate Treatment, NH_3 Strip
Sulfur Removal	CGCU-MDEA/CLAUS /SCOT (elemental sulfur)	CGCU-MDEA/CLAUS /SCOT (elemental sulfur)	CGCU-Selexol/CLAUS /SCOT (elemental sulfur)	CGCU-Selexol/CLAUS /SCOT (elemental sulfur)
Clean Fuel Gas/Gas Addition	Clean Fuel Gas Saturator (H_2O), N_2 Recycle from ASU	Clean Fuel Gas Saturator (H_2O), N_2 Recycle from ASU	Clean Fuel Gas Saturator (H_2O), N_2 Recycle from ASU	Clean Fuel Gas Saturator (H_2O), N_2 Recycle from ASU
Gas Turbine	Modified W501G	Modified W501G	Modified W501G	Modified W501G
Power (MWe)	275	266	258	258
PR/TIT (F)	17.67/2583	17.67/2583	17.67/2583	17.67/2583
Steam Cycle	3 Pressure	3 Pressure	3 Pressure	3 Pressure
Turb Pres	Level/Reheat	Level/Reheat	Level/Reheat	Level/Reheat
HP/IP/LP	1800/342/35 (psia)	1800/342/35 (psia)	1800/342/35 (psia)	1800/342/35 (psia)
Superheat/Reheat	1050/1050 F	1050/1050 F	1050/1050 F	1050/1050 F
Exhaust LP Turb	0.67 psia	2.89 psia	2.89 psia	1.2 psia
HRSG Stack Temp	260 F	260 F	260 F	260 F
Water Gas Shift			Two-Stage WGS	Two-Stage WGS
CO2 Sequestration				
Method	None	None	Selexol Dbl. Stage	Selexol Dbl. Stage
Delivery Option	None	None	HP L-CO_2 Stream	HP L- CO_2 Stream

Figure 5. Gasifier Layout

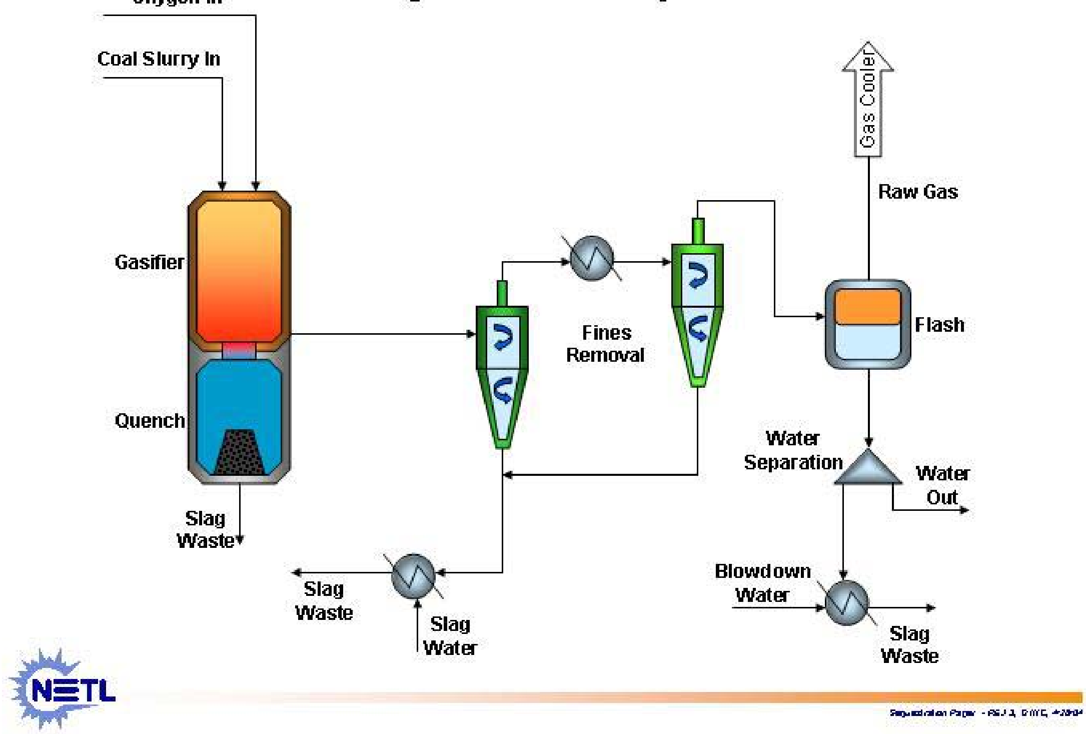


Table 2.3 Gasifier Conditions

	Case #1	Case #2	Case #3	Case #4
Coal (dry, T/d)	3460.50	3463.91	3541.23	3541.23
Coal (T/d)	4971.98	4976.88	5087.98	5087.98
Slurry Water (T/d)	236.20	236.43	241.71	241.71
Gasifier Pres (psia)	615	615	615	615
Gasifier Temp (°F)	2500	2500	2500	2500
Raw Fuel Gas Temp (°F)				
Quench Exit	425.6	425.6	424.3	424.3
To Gas Cooling	423	423	423	423
Heating Value (Btu/lb mol)				
LHV	39156	39058	39080	39080
HHV	52903	52819	52842	52842
Flow Rates (lb/hr)				
Coal Slurry	434,016	434,442	444,140	444,140
Oxidant (95% O_2)	268,319	268,570	274,249	274,249
Solid Waste Slurry	51,626	51,677	52,830	52,830
Water Purge	140,068	140,431	143,827	143,827
Makeup Water	148,618	149,470	95,228	95,228

2.3 Gas Cooling/Heat Recovery/Hydrolysis/Gas Saturation

The gas cooler layout is shown in Figure 6. The raw fuel gas from the gas scrubber is cooled in a series of heat exchangers to 103 °F and sent to the CGCU section. Any hydrogen chloride and ammonia is assumed to be in the condensate from these heat exchangers, which is then sent to an ammonia strip unit for further treatment. This section also contains a catalytic hydrolyzer in which the carbonyl sulfide is converted to hydrogen sulfide. Heat recovered in the heat exchanger network is used to generate low-pressure steam for

Table 2.4 ASU Summary

	Case #1	Case #2	Case #3	Case #4
% Air from GT	50	50	50	50
Air Inlet Pres (psia)	277	277	277	277
Total Air Flowrate (lb/hr)	1,127,210	1,128,270	1,152,130	1,152,130
Oxidant Stream				
Flowrate (lb/hr)	268319	268570	274249	274249
Purity (mole% O_2)	95	95	95	95
ASU Pres (psia)	92	92	92	92
Boost Comp Pres (psia)	650	650	650	650
Nitrogen Stream				
Flowrate (lb/hr)	414645	415033	464621	464621
Purity (mole% N_2)	98.88	98.88	98.87	98.87
ASU Pres (psia)	91/265	91/265	91/265	91/265
Boost Comp Pres (psia)	333	333	333	333
% Recycled to GT	46.56	46.56	51.05	51.05
GT Recycle Temp (°F)	700	700	700	700
Power Requirements (MWe)				
Air Boost Comp	25.75	25.75	26.29	26.29
O_2 Boost Comp	7.91	7.90	8.07	8.07
N_2 Boost Comp	7.24	7.23	8.19	8.19
N_2 Expander	7.21	7.21	6.52	6.52

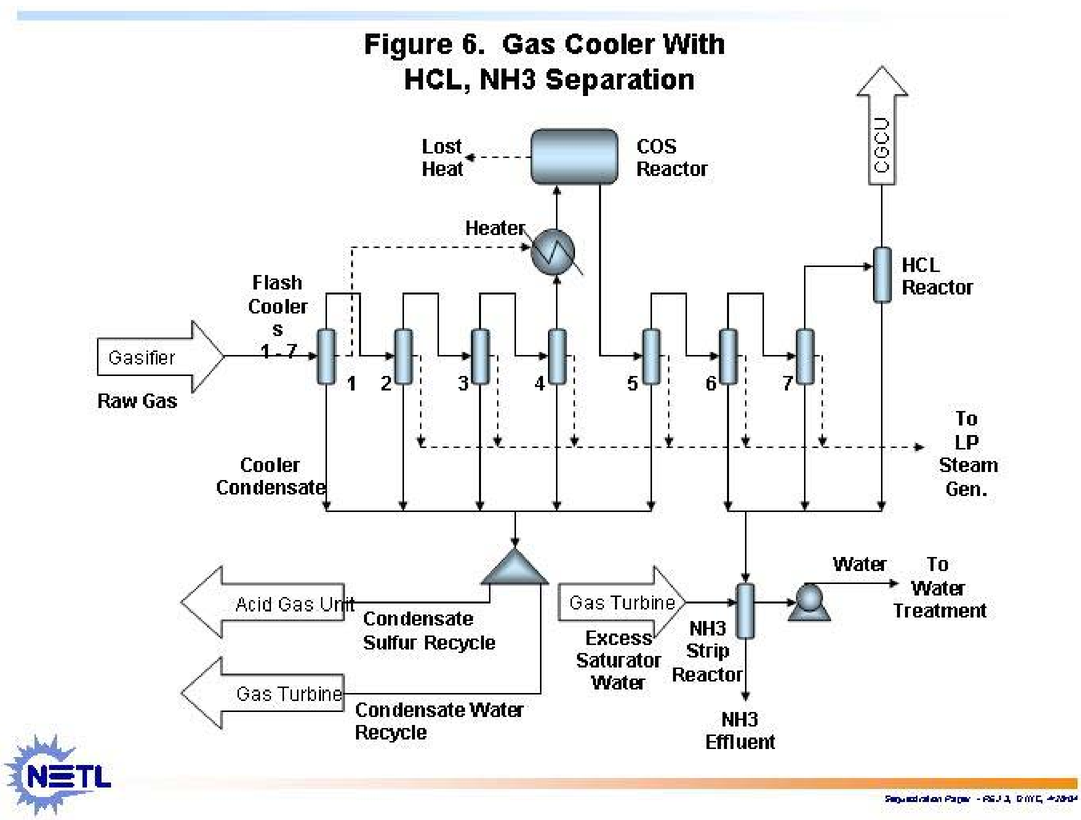


Figure 6. Gas Cooler With HCL, NH3 Separation

the HRSG and the ammonia strip unit. Additionally, low quality heat is used for BFW heating. The clean fuel gas from the CGCU is saturated with water using the high-pressure water condensate from the gas cooling unit before the fuel gas is sent to the gas turbine combustor. This lowers the amount of nitrogen recycle from the ASU needed to achieve the turbine power requirement to about 35%.

In cases #3 and #4, the MDEA unit is removed from the system, replaced by a series of two water-gas-shift (WGS) reactors and the first stage of a two-stage Selexol absorber unit, Figure 7. The WGS reactors are placed upstream of the gas cooling section. The WGS reactors are used primarily to shift the bulk of CO and any remaining methane to CO_2. The fuel gas stream enters stage one at 423 ^oF, and exits at 672 ^oF. Heat is removed between the two WGS stages (cooling the stream to 500 ^oF) and is sent to the steam cycle for steam generation. The second stage exhaust is approximately 594 ^oF. The WGS system also allows the replacement of the COS hydrolysis reactor.

2.4 Cold Gas Cleanup Unit (CGCU)

The MDEA/Claus/SCOT process is used for cold gas cleanup and sulfur recovery. Figures 8, 9, and 10 show the layouts for the MDEA unit, the Claus unit, and the SCOT process respectively. In the MDEA unit, the cooled gas from the low temperature heat recovery unit 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. 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 exits at 125 ^oF, 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.

As was mentioned earlier, in cases #3 and #4, the MDEA unit is removed from the system, replaced by the first stage of a two-stage Selexol absorber unit, shown in Figure 11. The first stage of the Selexol unit operates by introducing a CO2-rich stream of Selexol to the cooled fuel gas stream. The saturated stream selectively removes H2S from the fuel gas stream, leaving the CO2 and remaining fuel gas to pass to the second stage. The sulfur-laden stream is then sent to a stripper column where the H2S is removed and sent to the Claus unit, where the rest of the system is the same as in cases #1 and #2.

Further information about system performance is provided in Table 2.5. The sulfur recovery is improved by 1% with the change to the Selexol system. The sulfur content in the exit stream is of high purity (98-99%).

2.5 Gas Turbine

All cases were based on using a modified W501G gas turbine that was integrated with the ASU. A layout of the gas turbine and accessories is shown in Figure 12. From the compressor exhaust, a bleed stream is used to supply 50% 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. The remainder of the compressor discharge air is used to combust the

Figure 7. Gas Cooler, WGS, With HCL, NH3 Separation

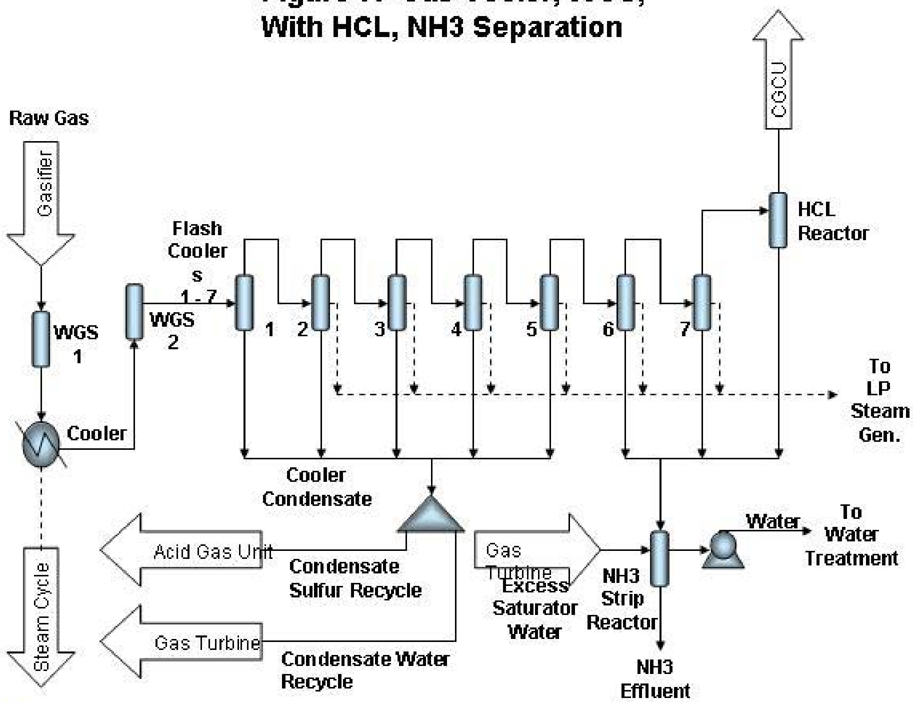


Figure 8. Acid Gas Unit

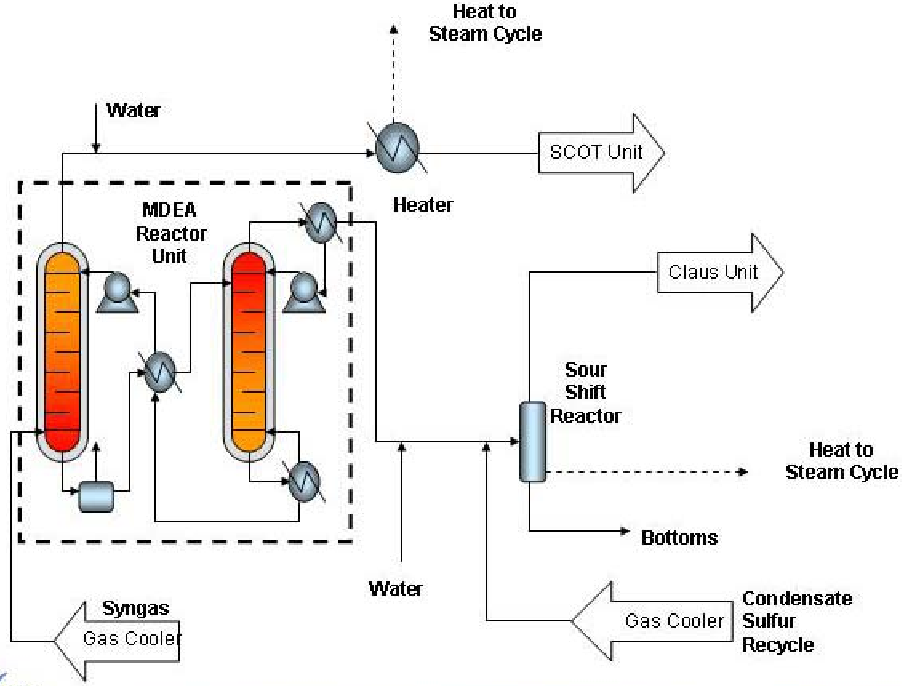


Figure 9. Claus Unit

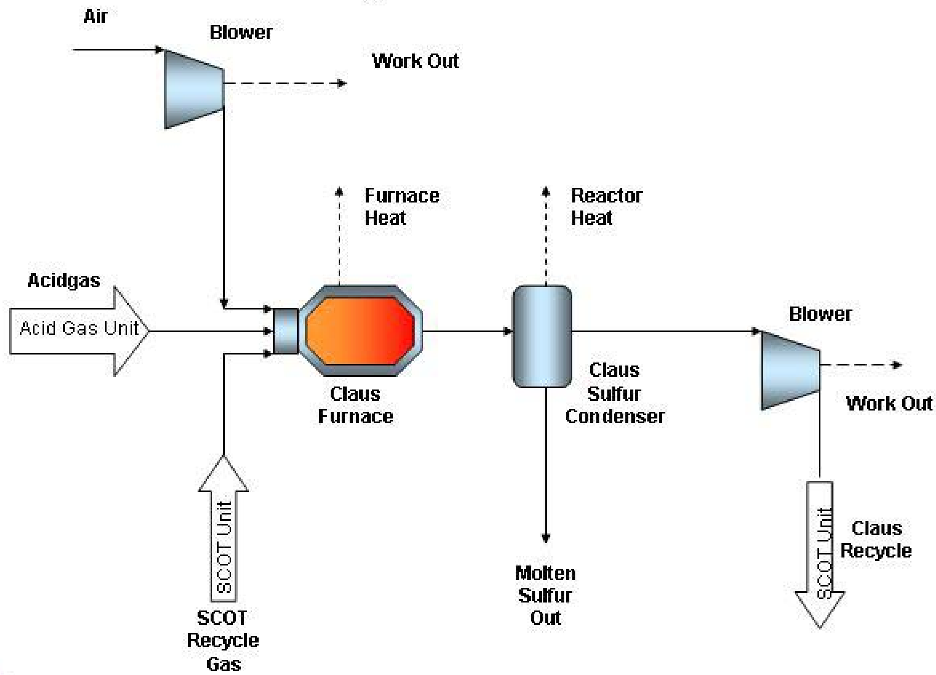


Figure 10. SCOT Reactor Layout

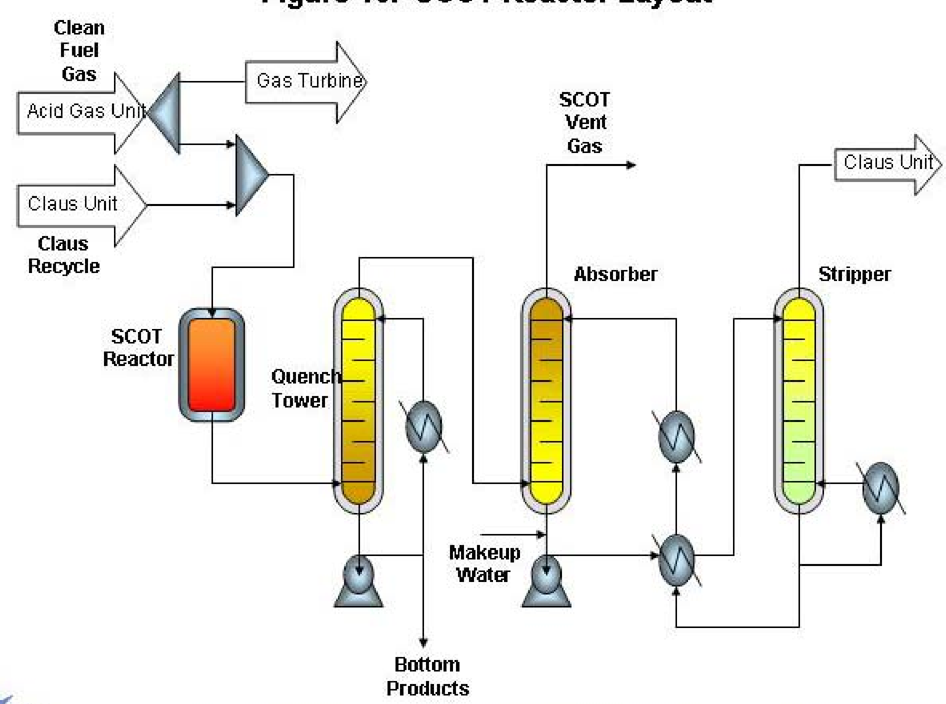


Figure 11. Acid Gas Unit

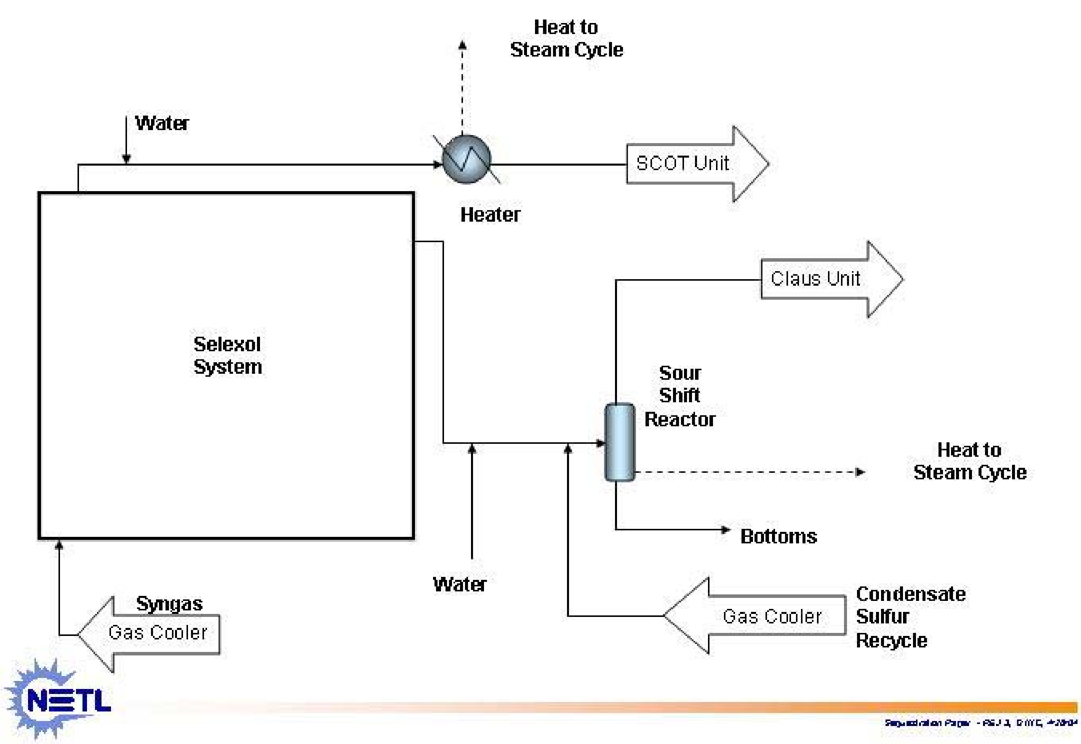
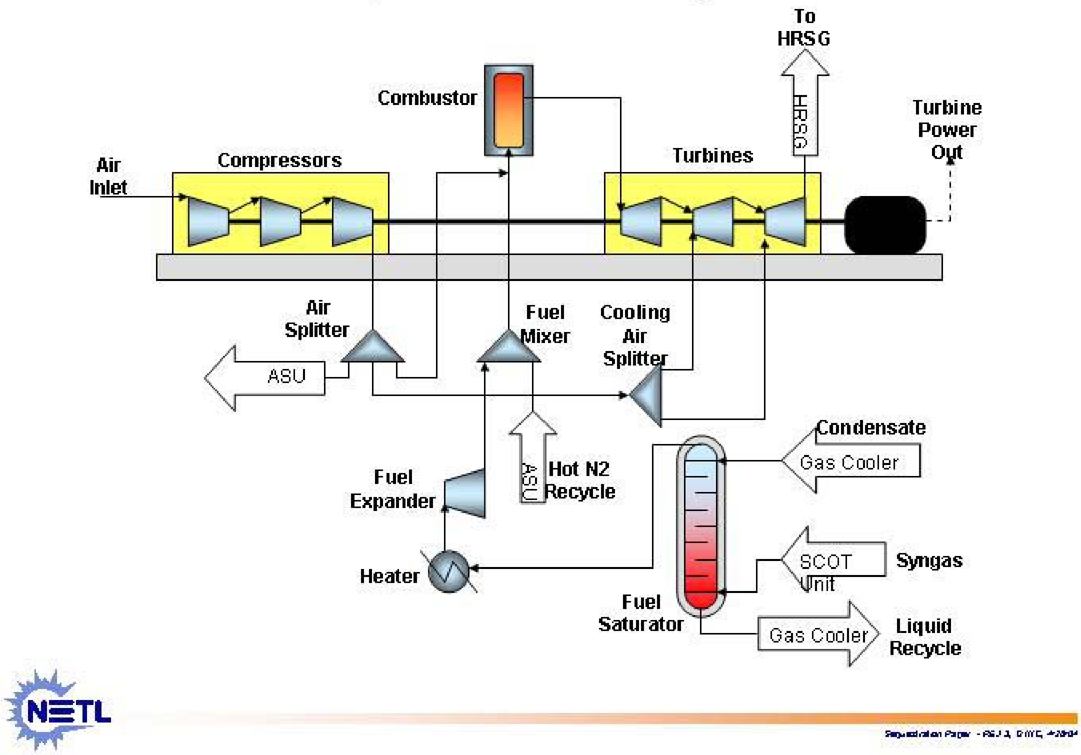


Table 2.5. CGCU Conditions

	Case #1	Case #2	Case #3	Case #4
Sulfur Balance (lb S/hr)				
Raw Gas	1929.588	1930.440	1963.767	1963.767
MDEA Feed	1888.632	1888.553	1942.104	1942.104
Saturator Recycle	33.436	34.33	13.86	13.86
Acidgas to Claus	1901.757	1902.57	1955.97	1955.97
Cleaned Fuel Gas	20.311	20.31	0.00	0.00
Sulfur Product	1899.238	1900.05	1953.19	1953.19
Fuel Recycle	0.288	0.287	0.000	0.000
SCOT Vent Gas	3.044	3.048	3.020	3.020
Key Conditions				
PPMV to CGCU	2293.01	2291.68	1585.9	1585.90
PPMV Clean Fuel Gas	25.64	25.62	0.00	0.00
Sulfur Recovery (weight%)	98.43	98.43	99.46	99.46
Steam requirements (lb/hr)	75364	75395.93	105790.00	105790.00
Power Requirements (kWe)	649.81	651.93	441.54	441.54

clean fuel gas. The ASU returns a nitrogen stream to the gas turbine combustor to assist in NOX control and to increase the flow rate and the power generated in the turbine expander. The nitrogen recycle flow rate is set by requiring that the gas turbine power generated equals approximately 274 MWe for the non-water-gas-shifted cases (#1 and #2), and 250 MWe for the shifted cases (#3, #4). Combustor duct cooling

Figure 12. Gas Turbine Layout



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 ^oF, 269 psia), where energy is recovered to produce power. The coal flow rate to the gasifier is adjusted up or down to allow the turbine inlet temperature to reach 2583 ^oF.

The original turbine design specifications are based on a natural gas fuel rather than a coal derived syngas. The syngas has a significantly lower heating value when compared to natural gas and requires a higher mass 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 low-BTU syngas. In the cases considered the syngas composition varies depending on the fuel processing prior to the gas turbine and the amount of nitrogen recycled from the ASU. In Table 2.6, the fuel gas composition for each case is listed both with and without the nitrogen stream addition. In Table 2.7, the gas turbine conditions are listed.

2.6 Steam Cycle

A schematic of the steam turbine layout is shown in Figure 13. 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. The gasifiers quench design results in no high quality heat being available for generating high pressure steam from the raw fuel gas. The system uses CGCU, but the higher gasifier pressure used results in differences in the low quality heat recovery sections. Sufficient heat is provided for reheating the condensate from the steam condenser and for the ammonia stripping unit. The higher pressure has heat of sufficient quality (i.e. high enough temperature) to be used for generating low pressure steam for use in the CGCU section and for use in the low pressure steam turbine section. A bleed of high pressure boiler feed

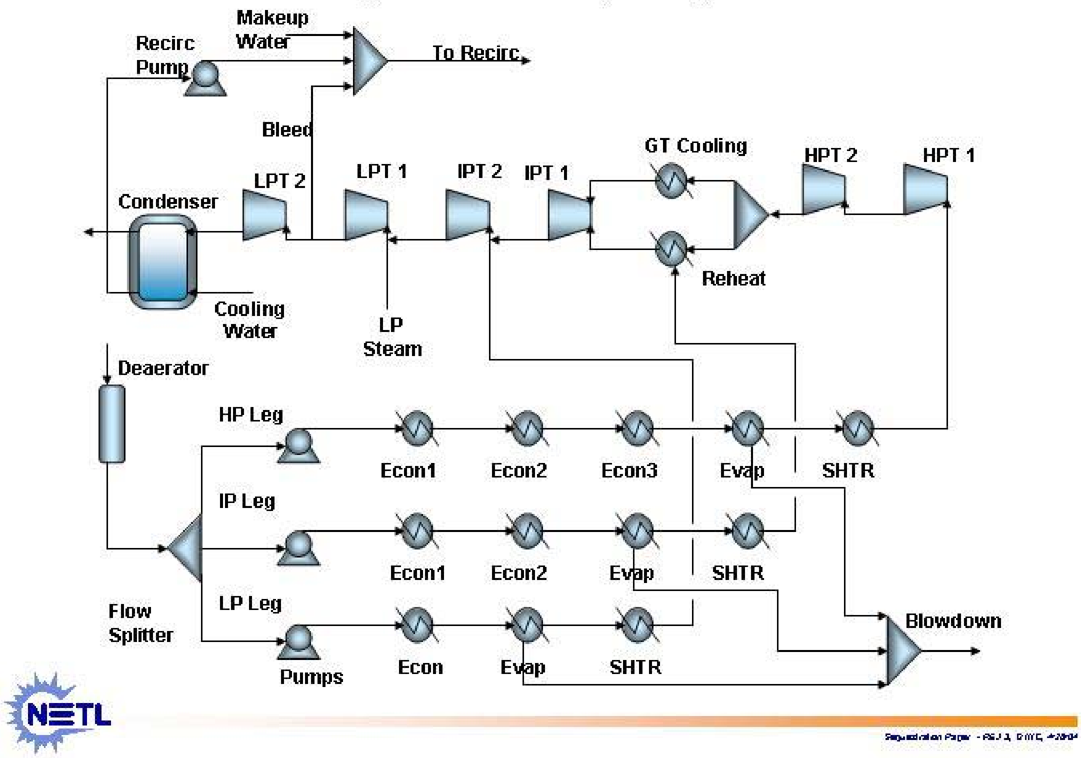
Table 2.6. Fuel Gas Composition

	No Nitrogen Recycle				Nitrogen Recycle			
	Case #1	Case #2	Case #3	Case #4	Case #1	Case #2	Case #3	Case #4
Mole %								
O_2	0.00	0.00	0.00	0.00	0.21	0.21	0.25	0.25
N_2	0.82	0.81	1.05	1.05	33.84	33.82	41.49	41.49
Ar	0.79	0.79	1.04	1.04	0.65	0.65	0.77	0.77
H_2	29.77	29.74	86.64	86.64	19.75	19.73	50.82	50.82
CO	39.79	39.75	0.00	0.00	26.39	26.37	0.00	0.00
CO_2	12.32	12.30	2.10	2.10	8.18	8.17	1.25	1.25
H_2O	16.42	16.52	9.15	9.15	10.92	10.99	5.41	5.41
CH_4	0.06	0.06	0.00	0.00	0.04	0.04	0.00	0.00
PPM								
H_2S	16.51	16.48	0.00	0.00	10.95	10.93	0.00	0.00
COS	4.69	4.68	0.00	0.00	3.11	3.10	0.00	0.00
NH_3	236.08	237.78	147.81	147.81	156.57	157.74	86.71	86.71
HCL	1.30	1.40	0.66	0.66	0.86	0.93	0.39	0.39
Heating Value HHV – Btu/lbm	4273	4270	21505	21505	2530	2529	4360	4360

Table 2.7. W501G Gas Turbine Conditions

	Case #1	Case #2	Case #3	Case #4
Pressure (psia)				
- to Filter	14.7	14.7	14.7	14.7
- Compressor Inlet	14.5736	14.5736	14.5736	14.5736
- Compressor Outlet	282.24	282.24	282.24	282.24
- Combustor Exit	268.52	268.52	268.52	268.52
- Expander Exhaust	15.2	15.2	15.2	15.2
Pressure Ratio	17.67	17.67	17.67	17.67
Flowrates (lb/hr)				
- Compr Inlet Air	4,320,000	4,320,000	4,320,000	4,320,000
- Fuel Gas	601,470	602,427	118,152	118,152
- Nitrogen Recycle	414,645	415,033	464,621	464,621
- Bleed Air to ASU	564,318	564,846	576,790	576,790
- Air Cooling Bleed	527,109	527,109	527,109	527,109
- Air Compr Leakage	13,478	13,478	13,478	13,478
- Steam Combustor Duct Cooling	70,000	70,000	70,000	70,000
- Expander Exhaust Gas to HRSG	4,758,320	4,759,130	4,312,510	4,312,510
Temperature (^oF)				
- Inlet Air	95	95	95	95
- Compressor Outlet	894.2883	894.8889	894.8889	894.8889
- Nitrogen Recycle	700	700	700	700
- Fuel Gas	466.4275	466.2001	460.7573	460.7573
- Combustor Exhaust	2612.612	2612.626	2612.818	2612.818
- Turbine Inlet	2583.293	2583.319	2582.947	2582.947
- Turbine Exhaust	1138.442	1138.583	1091.714	1091.714
Power (MWe)				
- Compressor	253.22	253.15	253.15	253.15
- Expander	529.42	529.59	507.54	507.54
- Generator Loss	3.87	3.87	3.56	3.56
- Net Gas Turbine	272.33	272.57	250.83	250.83
- Fuel Expander	5.40	5.39	4.35	4.35

Figure 13. Steam Cycle Layout



water is used for reheating the clean fuel gas from the CGCU section. This was the only convenient means for this case. The cooled boiler feedwater is re-pumped to the HRSG.

The primary heat recovered is from the exhaust gas stream of the gas turbine and the syngas coolers. Additionally, heat is integrated from the gas turbine cooling air chiller, from cooling the gasifier fuel gas, 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 stream. High pressure BFW for reheating the fuel gas is extracted after the third high pressure economizer section. Steam for the gas turbine combustor duct cooling is exctracted 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 ^{o}F and 0.67 psia for case #1. The steam cycle conditions are summarized in Table 2.8.

Case #2 and #3 see the replacement of the once-through condenser cooler with a direct air-cooled condenser, shown in Figure 14. Due to the higher ambient temperature of the Southwest US, the ability of the condenser to achieve the same approach temperature is lost, leading to a higher condensate temperature and higher pressure on the back-side of the steam turbine. This robs the system of a considerable amount of power (23 MWe). In case #4, a wet tower condenser is used to replace the direct air-cooled condenser, with saline aquifer water as the working fluid, shown in Figure 15. The system uses a mechanical draft cooling tower to cool the recycle water. Concentration ratio for the salts in the recycle water is set at 2:1. The water is brought out of the ground via deep wells, and the blowdown of high concentration brine is sent down the well hole upstream of the takeout point.

Table 2.8. Steam Cycle Conditions

HRSG Stack Gas Temperature				261		
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 ^{o}F		
Intermediate Pressure Turbine Inlet				342 psia/1050 ^{o}F		
Low Pressure Turbine Inlet				35 psia		
Low Pressure Turbine Exhaust				0.67 psia (Case#1) 2.89 psia (Case #2, #3) 1.19 psia (Case #4, #5)		
	Steam Conditions		**HRSG Approach Delta Temp (^{o}F)**			
Pressure Level	**Pressure (psia)**	**Saturation Temp. (^{o}F)**	**Case #1**	**Case #2**	**Case #3**	**Case #4**
Low	72.5	305	45.51	45.53	61.28	61.28
Intermediate	352	432	26.59	26.63	60.77	60.77
High	1911	629	60.45	60.43	61.01	61.01
Power Production (MWe)	**Case #1**	**Case #2**	**Case #3**	**Case #4**		
Steam Turbines	159.538	136.943	142.067	158.247		
Generator Loss	2.393	2.054	2.131	2.374		
Net Steam Turbines	157.145	134.889	139.936	155.874		
Pumps	1.624	1.623	1.397	1.399		
Air Compressor -Condensers	0.000	8.799	8.799	4.308		
Water Pumps (Tower)	0.000	0.000	0.000	1.898		

Figure 14. Steam Cycle Layout

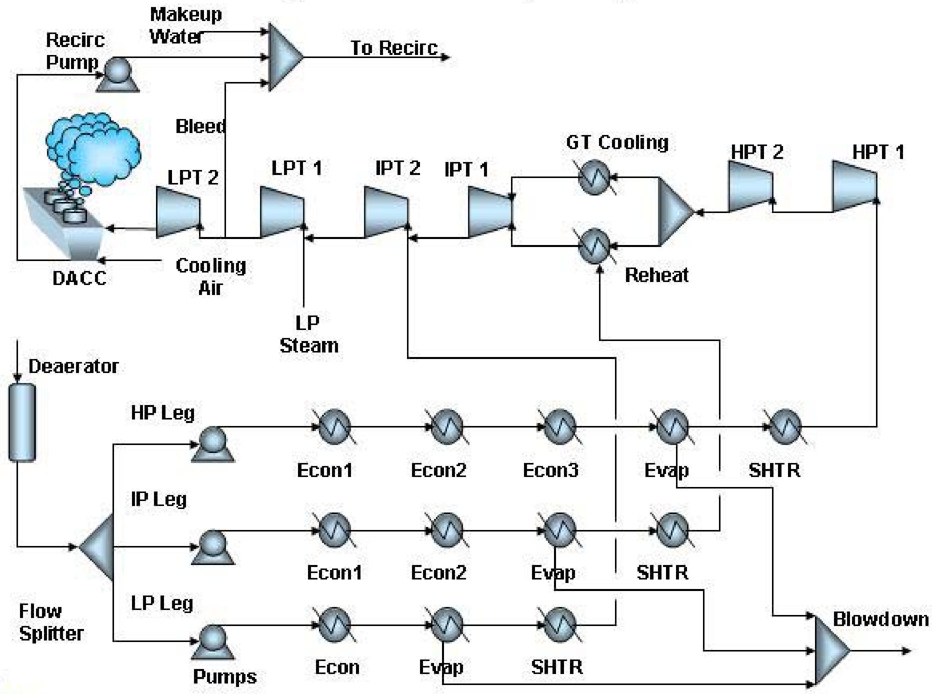
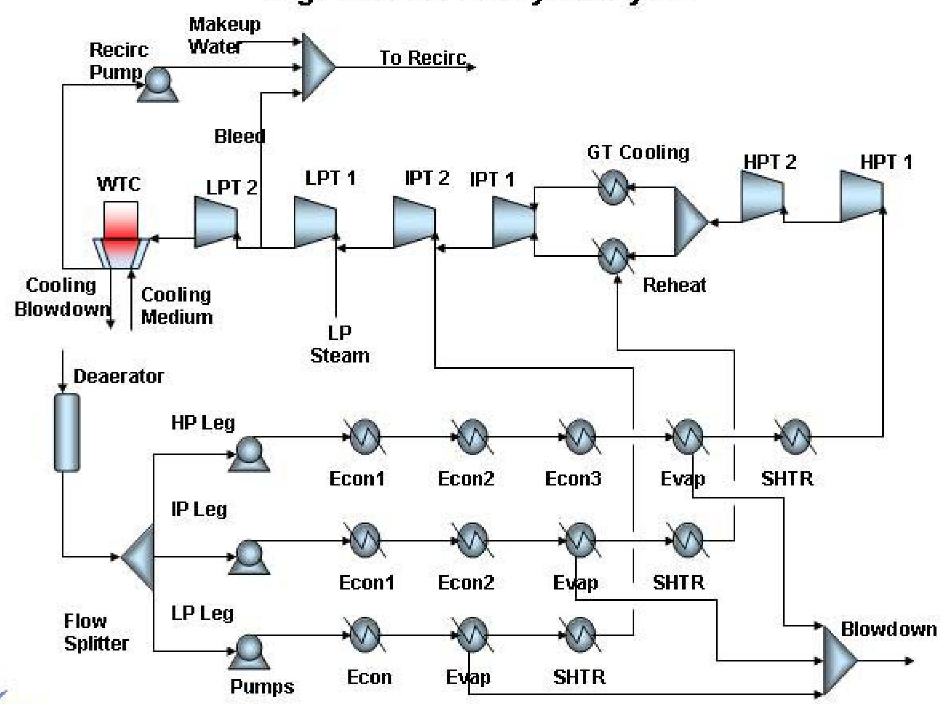


Figure 15. Steam Cycle Layout



2.7 CO2 Sequestration

CO_2 sequestration was introduced into the system for cases #3 and #4. The purpose of the sequestration system was to remove CO_2 from the fuel syngas stream. A Selexol two-stage unit was chosen for this operation, shown in Figure 16. The unit's first stage selectively removes sulfur from the fuel syngas, as was mentioned in section 2.4. The second stage of the Selexol unit selectively removes CO_2 from the clean syngas stream. The unit removes 97% of CO_2 in the syngas stream. The offgas of the second stage, cleaned of sulfur and scrubbed of CO_2, continues into the combustor section of the GT.

The CO_2 which has been removed from the syngas stream is cooled and compressed via a multi-stage compressor. The CO_2 is compressed and cooled to a 2100 psia, liquid CO_2 stream. Compressor power consumed in cases #3 and #4 is 26.7 MWe. The CO_2 stream, after compression, is sent for injection into the local glacial saline aquifer for sequestration. The sequestered stream has virtually no water content, with 96% CO_2 content. Primary contaminants in the stream are: 2.5% carbon monoxide, 1.2% hydrogen, and 0.1% methane.

2.8 Power Production

An auxiliary power consumption is assumed as 3 percent of the total power production by the gas turbine and steam turbine minus the power consumed by the miscellaneous pumps, expanders, compressors, and blowers. The power production and the overall process efficiency are listed in Table 2.9.

Figure 16 Two-Stage Selexol System

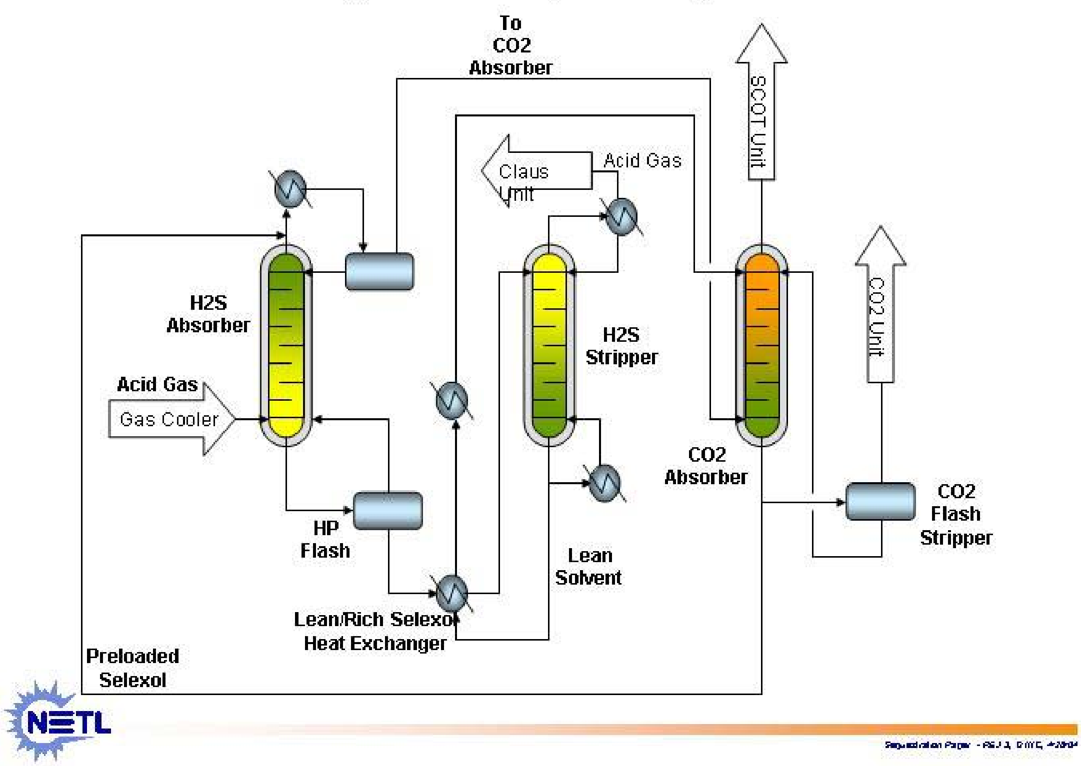


Table 2.9. Power Production

	Case #1	Case #2	Case #3	Case #4
Gas Turbine (MWe)	272.33	272.57	250.83	250.83
Steam Turbine (MWe)	157.14	134.89	139.94	155.87
Miscellaneous (MWe)	31.21	39.99	69.49	66.90
Auxiliary (MWe)	11.95	11.02	9.64	10.19
Plant Total (MWe)	386.32	356.44	311.64	329.61
Efficiency (HHV, %)	38.89	35.85	30.66	32.42
Efficiency (LHV, %)	40.45	37.28	31.86	33.70

3.0 Economic Analysis and Results

Costs of components and the overall system were developed based on 1st quarter 1999 costs (latest information available at the time the report was developed). Plant sections are broken out in the various tables in Chapter 3. A project contingency of 15% was added to all components. A variable process contingency was applied to each plant section based upon the maturity of the technology being costed, leading to an overall process contingency factor of 3 - 4%. Conceptual equipment was handled by being compared to a similar components price, adjusting based upon manufacturing differences, and applying 15% process contingency due to uncertainties in the cost.

3.1 Component Costs

The cost of electricity for the 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 and the OSPS Quality Guidelines for System Studies (1) document. Details of the individual section costs are described below and are based on capacity-factored techniques. The COE spreadsheets are included at the end of the report. All costs are reported in 1st Quarter 1999 dollars. Table 3.1 shows the major equipment cost comparisons

Table 3.1. Major Equipment Cost Comparison

($Million)	Case #1	Case #2	Case #3	Case #4
Coal Slurry Preparation	$36,163	$36,188	$36,752	$36,752
Oxygen Plant	$71,565	$71,598	$72,932	$72,932
Texaco Gasifier (Quench)	$44,464	$44,494	$45,187	$45,187
Low Temperature Gas Cooling/Gas Saturation	$12,384	$12,406	$7,595	$7,595
MDEA	$5,753	$5,755	$0	$0
Claus	$5,888	$5,889	$5,955	$5,955
SCOT	$9,511	$9,514	$5,079	$5,079
Selexol	$0	$0	$49,088	$49,088
Gas Turbine System	$54,041	$54,090	$49,776	$49,776
HRSG/Steam Turbine	$46,561	$43,538	$42,790	$44,951
Condenser	$0	$15,866	$15,809	$6,039

The coal slurry preparation section includes costs for coal hoppers, feeders, conveyors, sampling and feed systems. The cost for the oxygen plant includes the air separation unit, the air precoolers, the oxygen compressors, the nitrogen compressors and the air compressors. The cost for the gasifier was derived from a previous Texaco report and is dependent on the cooling process used within the gasifier.

The cost for the low temperature cooling and gas saturation section includes several heat exchangers, separators, the saturator, fuel gas reheaters, and the turbine expander. The cost of the MDEA acid gas removal system includes the absorber column, the stripping column, heat exchanger and pumps. The cost for the Claus/SCOT sulfur recovery and tail gas treating units is based on 88 tons per day of sulfur entering the unit. The selexol system is a double-stage absorption system. The cost of the selexol unit is based on a generic selexol unit of the same throughput. A 10% process contingency and a 15% project contingency were applied to account for any discrepancies.

The cost for the W501G gas turbine was derived from the Gas Turbine World 2001-2002 Handbook (2). 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. A process contingency of 5% was added to the total plant cost based on the development of modified gas turbines. The cost for the steam cycle is based on a three-pressure level steam cycle. Steam turbine power is 157.14 MWe for the base case. The cost of the condenser in case #1 is included with the cost of the steam turbine section. The costs of the Direct Air Cooled Condenser and the Wet Tower Condenser were calculated separate of the steam turbine section, and were calculated using the surface area requirement for the condensers.

3.2 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. Table 3.2 shows the percentages used in this report.

Table 3.2. Bulk Plant Percentages

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

3.3 Economic Performance

Table 3.3 shows the allowances for capital cost expenditures for the system. Table 3.4 shows allowances for O&M costs. Table 3.5 shows the investment assumptions used in the COE analysis. Table 3.6 shows total plant investment requirement for all cases. Table 3.7 shows the operating costs for all cases. Table 3.8 shows the levelized cost of electricity in current and constant dollars for all cases.

Table 3.3. Capital Cost Assumptions

Engineering Fee	10% of Process Plant Cost
Project Contingency	15% of Process Plant Cost
Construction Period	4 Years
Inflation Rate	3%
Discount Rate	11.2%
Prepaid Royalties	0.5% of Process Plant Cost
Catalyst and Chemical Inventory	30 Days
Spare Parts	0.5% of Total Plant Cost
Land	200 Acres @ $6500/Acre
Start-Up Costs	
Plant Modifications	2% of Total Plant Investment
Operating Costs	30 Days
Fuel Costs	7.5 Days
Working Capital	
Coal	30 Days
By-Product Inventory	30 Days
O&M Costs	30 Days

Table 3.4. Operating & Maintenance Assumptions

Consumable Material Prices	
Western Sub-Bituminous Coal	\$24.20 / Ton
Raw Water	\$5.00 / kgal
MDEA Solvent	\$1.45 / Lb
Claus Catalyst	\$470 / Ton
Shift Catalyst	\$5.00 / Lb
Selexol Sorbent	\$0.05 / Ton CO_2
SCOT Activated Alumina	\$0.067 / Lb
Sorbent	\$6000 / Ton
Nahcolite	\$275 / Ton
Off-Site Ash/Sorbent Disposal Costs	\$8 / Ton
Operating Royalties	1% of Fuel Cost
Operator Labor	\$34 / hour
Number of Shifts for Continuous Operation	4.2
Supervision and Clerical Labor	30% of O&M Labor
Maintenance Costs	2.2% of Total Plant Cost
Insurance and Local Taxes	2% of Total Plant Cost
Miscellaneous Operating Costs	10% of O&M Labor
Capacity Factor	85%

Table 3.5. 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%	9% Cost	7.2% Return
Preferred Stock	0%	3% Cost	0% Return
Common Stock	20%	20% Cost	4% Return
Total			11.2% Total
Book Life		20 Years	
Tax Life		20 Years	
State and Federal Tax Rate		38%	
Tax Depreciation Method		ACRS	
Investment Tax Credit		0%	
Number of Years Levelized Cost		10 Years	

Table 3.6 Total Plant Investment (in $1000's)

	Case #1	Case #2	Case #3	Case #4
Process Plant Cost	**$383,685**	**$401,112**	**$445,586**	**$437,176**
Engineering	$38,368	$40,111	$44,559	$43,718
Process Cont.	$11,749	$13,381	$18,189	$17,328
Project Cont.	$57,553	$60,167	$66,838	$65,577
Total Plant Cost (TPC)	**$491,354**	**$514,770**	**$575,172**	**$563,802**
Interest/Inflation	$61,679	$64,619	$72,201	$72,201
Total Plant Investment	**$553,034**	**$579,389**	**$647,373**	**$634,576**
Prepaid Royalties	$1,918	$2,006	$2,228	$2,186
Initial Catalyst/Chemical Inventory	$119	$120	$576	$576
Startup Costs	$13,714	$14,291	$15,769	$15,489
Spare Parts	$2,457	$2,574	$2,876	$2,819
Working Capital	$4,692	$4,718	$4,825	$4,815
Land	$1,300	$1,300	$1,300	$1,300
Total Capital Requirement	**$577,234**	**$604,398**	**$674,946**	**$661,760**
$/kW	**1494**	**1696**	**$2166/kW**	**$2008/kW**

Table 3.7 Annual Operating Costs (in $1000's)

	Case #1	Case #2	Case #3	Case #4
Coal	$37,330	$37,330	$38,201	$38,201
Water	$777	$777	$498	$498
MDEA	$181	$181	$0	$0
Claus	$1	$1	$1	$1
Shift Catalyst	$0	$0	$9	$9
Selexol	$0	$0	$129	$129
Scot Alumina	$3	$3	$3	$3
Scot Cobalt	$5	$5	$5	$5
Scot Chemicals	$16	$16	$16	$16
Ash Disposal	$846	$846	$865	$865
Operator Labor	$4,455	$4,455	$4,455	$4,455
Supervision	$2,634	$2,634	$2,855	$2,825
Maintenance	$10,810	$10,810	$12,654	$12,404
Royalties	$373	$373	$382	$382
Other	$878	$878	$952	$942
Total Operating Cost	**$58,310**	**$58,310**	**$61,025**	**$60,735**
By-Products				
Sulfur	$530	$530	$545	$545
Water	$0	$0	$0	$0
Net Operating Costs	**$57,780**	**$57,780**	**$60,480**	**$60,190**

Table 3.8 Cost of Electricity

Mills/kWhr	Case #1	Case #2	Case #3	Case #4
Current $'s	58.7	65.7	81.6	76.1
Constant $'s	49.6	55.5	68.8	64.2

In looking at the TPI numbers, cases with a change to DACC and then addition of CO_2 sequestration both incur capital cost increases. Switching to the Wet Tower Cooling setup shows a small improvement in costs, whereas the addition of the hydrate system has no affect on the plant capital cost.

Looking at the operating costs, there is no influence due to addition of the DACC system. Addition of the CO_2 sequestration system shows an increase in operating costs. Wet cooling, like the DACC, has no major affect on the operating costs. The COE numbers follow the same trends as the TPI results.

4.0 Conclusions

The finished study came to the following case by case conclusions.

Case #2 –

- Dry air cooling adversely impacted the system performance and economics when compared to once-through wet cooling technology.
  - Loss of 17 MW of net steam turbine power.
  - Loss of 30 MW in overall net power from the plant.
  - Increase of $13 million in bare equipment costs.
  - Increase of $27 million in TPI, and ~$200/kW.
  - No impact on operating costs.
  - Increase in COE by 12.5%

Case #3 –

- CO_2 sequestration further impacted the system performance and economics.
  - Decrease in primary fuel mass flow to the gas turbine, forcing an increase in nitrogen recycle from the ASU for makeup mass flow. Turbine output, however, still drops by 22 MW due to changes in vane arrangement to account for switching from syngas to primarily hydrogen in the water-gas-shifted syngas.
  - Increase of 5 MW net power in the steam turbine system. It is not known at this time what causes this increase.
  - Loss of 45 MW in overall net power from the plant.
  - Increase of $49 million bare equipment cost for the Selexol system.
  - Increase of $70 million in TPI, and ~$370/kW
  - Small increase in the operating costs of the plant.
  - Increases in the COE of 20%.

Case #4 –

- Switching to wet tower cooling minimizes some of the impact seen by the two previous technological advances, lowering costs and improving efficiency.
  - 16 MW net gain in steam turbine power over a DACC system, with CO_2 sequestration.
  - Saves approximately 2.6 MW worth of auxiliary power loss over the DACC system.
  - Improvement of 18 MW overall net power for the plant.
  - Savings of $8.5 million in bare equipment costs.
  - Savings of $13 million in TPI, and ~$160/kW.
  - No affect on operating costs.
  - Small decrease in COE.

Overall, conclusions generated from the study show that DACC is shown to be an ineffective answer to cooling problems for water starved regions. Also, the CO_2 sequestration system delivers a high purity liquid CO2 stream for either geologic sequestration or across-the-fence delivery. Moreover, CO_2 sequestration, though environmentally beneficial, is shown to increase costs of electricity production. Addition of WTC is an effective way to cool the Steam condenser, and offset some costs associated with CO_2 sequestration

5.0 References

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