Appendix B
Case 2: Lurgi gasifier

CASE 2

The following report gives a brief description of each of the units in the block flow diagram. All capital cost data in this report, except where otherwise specified, has been estimated from similar installations described in the Houston Area Medium-BTU Coal Gasification Project Final Report, published in June 1982 by Union Carbide [1] (All references to material in this report will be referred to as Houston, and all scaling exponents from the Houston report are 0.65). The plant consumes 5.0 million metric tons of coal, 2.1 million metric tons of oxygen, and produces 0.50 million metric tons of mixed alcohols per year.

COAL PREPARATION

Coal (stream 10) is sent to the Coal Preparation Block. The coal is crushed and conveyed to the lockhopper on top of the gasifier. The Coal Preparation Block is composed of three plants from the Houston report. Plant 61 is the Reclaiming, Transfer, and Crushing Plant. The cost of this plant was scaled exponentially. Plant 22 is the Barge Terminal. This plant was scaled exponentially. Plant 60 is Coal Receiving and Storage and again the cost for this plant was scaled exponentially.

LURGI GASIFIER

The coal enters the Lurgi gasifier through a lockhopper at the top. The coal flows down in the gasifier, counter-current to the gases. The ash settles to the conical bottom of the gasifier. The compressed oxygen (stream 9) and steam (stream 12) enter beneath the surface of the ash bed, which is supported by a rotating grate. Ash exits from the bottom of the gasifier. The syngas exits near the top of the gasifier at 400°C and 2,800 kPa. The raw syngas is quenched, some heat is recovered, and the heavy by-products condense and are removed. Economic data for this block was obtained from a report on the Great Plains Gasification Project, and the cost for this plant was estimated by using exponential scaling, where each train can handle up to 958.3 tons of coal per day [5].

CRYOGENIC OXYGEN PLANT

Compressed air (stream 1) is cooled and sent to the Cryogenic Oxygen Plant Block, and is separated into high purity oxygen (stream 2), nitrogen (stream 3), argon (stream 6), and a water and carbon dioxide waste mixture (stream 28). A small quantity of nitrogen (stream 19) is sent to the Rectisol Block. The Cryogenic Oxygen Plant Block does not include the inlet air compressors or the outlet oxygen compressors. In the cryogenic system, there are provisions for gaseous and liquid oxygen backups sufficient to maintain downstream plant operation in the event of a shutdown in the cryogenic facility. The capital investment has been calculated linearly for the reduction in trains, and exponentially for throughput change per train. Each train can produce up to 2,000 tons of oxygen per day. The Houston plants that comprise the Cryoplant Block are 02 and 08.

RECTISOL

The cooled raw gas stream (stream 18), nitrogen gas (stream 19) for methanol regeneration, and methanol make-up (stream 20) for vapor loss all enter the Rectisol Block. H_2S levels are reduced to the ppb range and CO_2 levels to the ppm range. The clean syngas (stream 22) is sent to the alcohol synthesis loop. A CO_2-N_2 mixture (stream 24) and a CO_2 rich stream (stream 23) are produced as byproducts. Condensed water is also removed (stream 17A). This block is the same as Houston Plant 05. The cost for this plant was estimated by using exponential scaling.

COS HYDROLYSIS

The sulfide rich stream from the Rectisol Block (stream 25) and steam are sent to the COS Hydrolysis Block where COS is converted to H_2S. The product gas (stream 41) is sent to the Claus Sulfur Recovery Block. The COS Hydrolysis Block cost is assumed to be negligible.

CLAUS PLANT

Hydrogen sulfide rich gas (stream 25) is mixed with air (stream 42) and converted in a two-step reaction to elemental sulfur (stream 46). The unreacted hydrogen sulfide (stream 45) is then sent to the Beavon Plant for further treatment. This block is the same as Houston Plant 06. The cost for this plant was estimated by exponential scaling.

BEAVON PLANT

The Claus tail gas (stream 45) and air (stream 47) go to the Beavon Block. Additional sulfur is made (stream 51), and the gas leaving (stream 50) is sufficiently free from sulfides that it can be vented to the atmosphere. A sour water stream (stream 54) is sent from the plant for treatment. The cost of this block was estimated from data collected from various sources, with a scaling exponent of 0.65 [2].

MoS_2 ALCOHOL SYNTHESIS LOOP

Clean syngas (stream 26) at 140 atmospheres enters the catalytic reactor along with the syngas recycle (stream 56B). The products (stream 26A) are taken to the separations block where the unreacted syngas is removed (stream 59). Part of this stream (stream 27) is sent to power generation while the rest (stream 56) is sent to CO_2 removal. The cost of this block was estimated from the cost of a methanol synthesis loop, with a scaling exponent of 0.565 [3].

CO_2 REMOVAL

This block is very similar to the Rectisol Block. Recycled gas from the alcohol separation block (stream 56) is the only feed. CO_2 free syngas (stream 56A) is then recompressed and sent back to the reactor. CO_2 is taken off as a product (stream 57). The cost of this block is calculated the same way as in the Rectisol block. The power requirement for this unit has been included in the Rectisol block.

COMBUSTION GAS TURBINE

The light hydrocarbons extracted from the reactor recycle (stream 27) in the Alcohol Synthesis Loop are sent to a combustion gas turbine with hot gas heat recovery. The power from the combustion gas turbines is assumed to be 35% of the HHV of the fuel in stream 27. This is consistent with recent studies on IGCC plants using medium BTU synthesis gas [7]. The cost for this block was estimated from data taken from an EPRI report, where each train can produce up to 200 MW with a scaling exponent of 0.67 [8].

EXHAUST GAS HEAT RECOVERY

The hot exhaust gas stream from the Gas Turbine Block (stream 70) at 590°C and 101 kPa enters the Exhaust Gas Heat Recovery Block and is cooled against process boiler feed water at 25°C (stream 73). The exhaust gas stream exits at 200°C (stream 75), and the boiler feed exits as steam at 10,000 kPa and 535°C (stream 74). The cost for this block was estimated from data taken from an EPRI report, where each train can generate up to 425 tons of steam per hour with a scaling exponent of 0.67 [8]. This block also supplies the reheat between the high pressure and intermediate pressure steam turbines.

POWER GENERATION

The steam from the Exhaust Gas Heat Recovery Block is let down in the steam turbines for power production. The cost for this block was estimated from data taken from an EPRI report, where each train can produce up to 500 MW with a scaling exponent of 0.67 [8]. This is a 3-stage steam turbine system. The high pressure stage inlet is 535°C, 10,000 kPa steam. The exhaust at 3,000 kPa is reheated to 535°C before entering the intermediate pressure stage. The final stage exhausts to a surface condenser at 7.4 kPa. Each turbine has an assumed efficiency of 75%.

PRESSURE SWING ADSORPTION

The clean syngas (stream 22A) is sent to PSA for selective hydrogen removal and purification. The adjusted syngas (stream 22AA), which leaves the PSA unit at 200 kPa, is compressed and sent on to the reactor while the purified excess hydrogen is sent to the plant battery limits for sale if possible.

IMPORTANT POINTS OF INFORMATION

Several decisions were made for the creation of this case that should be pointed out. Also, there are alternatives that have not been fully considered which should be viewed in more detail. They are listed below along with the reasons behind them.

- The traditional method of producing large quantities of pure oxygen is by cryogenics, which is used for this case. However, recent reports have suggested that membrane and catalytic processes are becoming economically competitive with cryogenics. Therefore, we suggest that more research be done in this area.

- The Rectisol system was chosen for this case for H_2S and CO_2 removal. The major alternative to Rectisol is Selexol. The literature indicates that Rectisol has a higher installed capital cost, but a lower fixed operating cost than Selexol. Both of these systems are capable of removing H_2S to the ppm level and beyond. However, there is some evidence that substantial quantities of H_2S are beneficial in the reaction stages of this case. If this is so, then a system such as the Benfield Acid Gas removal process might be more suitable. The Benfield system does not remove as much H_2S and has lower capital and operating costs.

- The operating pressure for the Lurgi Gasifiers has been set at 2,800 kPa. No data have been found on the maximum pressure that a Lurgi Gasifier can be run. Since the pressure required at the reactor is 14,000 kPa, we would of course like to run the gasifiers at as high a pressure as possible.

TOTAL ESTIMATED CAPITAL INVESTMENT (MM$)

Pressure Swing Adsorption	20.6
Coal Preparation	77.5
Lurgi Gasifier	653.4
Gas Turbines	356.4
Steam Turbines	46.8
Exhaust Gas Heat Recovery	29.6
Cryogenic Oxygen Production	212.7
Rectisol (Acid Gas Separation)	115.5
Claus (Sulfur Recovery)	29.7
Beavon	6.4
Alcohol Synthesis Loop	47.2
CO2 Removal	4.5
Other Compressors	216.6
TOTAL	1817.1

(sum of individual block costs does not exactly equal the total due to round-off)

OVERALL ECONOMIC EVALUATION

The following table gives the totals and breakdowns for the yearly operating costs as well as the total installed cost for the plant.

TOTAL ESTIMATED INSTALLED CAPITAL COST (MM$)		1817.1
TOTAL ESTIMATED OPERATING COSTS (MM$/YR)		483.7
Coal ($33/metric ton delivered)	163.5	
Other Expenses	320.2	
TOTAL ESTIMATED CREDITS (EXCLUDING ALCOHOLS) (MM$/YR)		430.2
Power ($0.05/kWh)	312.0	
Sulfur ($300/metric ton) (7)	35.2	
Coal Tar/Liquid By-Products ($99.1/metric ton)	43.4	
Hydrogen ($35.31/1000 cubic meters)	39.5	

Credits for nitrogen, argon, and other rare gases have not been included because prices were not available and potential markets have not yet been identified.

STAND ALONE COMPRESSORS AND POWER SUMMARY

There are 4 compressors that are not included in any of the blocks. Their inlet, outlet, pressure change, power rating, and installed capital cost are listed below. Following that is a summary of the total plant power output/input (4). An efficiency of 70% is assumed for all compressors, with a maximum pressure ratio of 5 for a single stage of compression. Multiple compression stages with intercooling are used for services with pressure ratios greater than 5.

FUNCTION	INLET STREAM	P (kPa)	OUTLET STREAM	P (kPa)	POWER (MW)	COST (MM$)
Air Prep	1A	101	1	500	-81.4	71.4
O2 Prep	2	500	9	2800	-16.0	15.0
Rxtr Prep	22AA	200	26	14000	-148.0	129.8
Recy Comp	56A	12666	56B	14000	-0.2	0.4
Total compressor needs					-245.6	
Other in plant needs					-39.5	
Total produced in steam and gas turbines					1065.1	
Net power output					780.1	
Total installed compressor costs (1992 dollars)						216.6

REFERENCES

1. *Final Report on the Houston Area Medium-BTU Coal Gasification Project, Volumes 2 and 3.* Prepared by the Linde Division of Union Carbide Corporation, June 1982.

2. "Beavon Sulfur Removal Process," *Hydrocarbon Processing*, April 1984, p. 78.

3. Frank, Marshall E. "Methanol: Emerging Uses, New Syntheses," *Chemtech*, June 1982, pp. 358-362.

4. Baasel, William D., *Preliminary Chemical Engineering Plant Design, 2nd edition,* Van Nostrand Reinhold, New York, 1990, pp. 529-530.

5. Miller, W. R., F. I. Honea, R. A. Lang, T. E. Berty, R. C. Delaney, R. W. Hospodarec, and P. F. Mako, "Great Plains Gasification Plant Start-Up and Modification Report," U.S. Dept. of Energy report no. DOE/CH/10088-2018, March 1986.

6. *Chemical Marketing Reporter*, August 31, 1992.

7. Report TR-101789, Houston Lighting and Power Company's Evaluation of Coal Gasification Coproduction Energy Facilities, EPRI Project 3226-04, 1992.

8. EPRI Report TR-100319, *Evaluation of a 510-MWe Destec GCC Power Plant Fueled With Illinois No. 6 Coal*, Prepared by Fluor Daniel, Inc., EPRI Project 2733-12, 1992.

Figure B.1 : Block Flow Diagram for Case 2

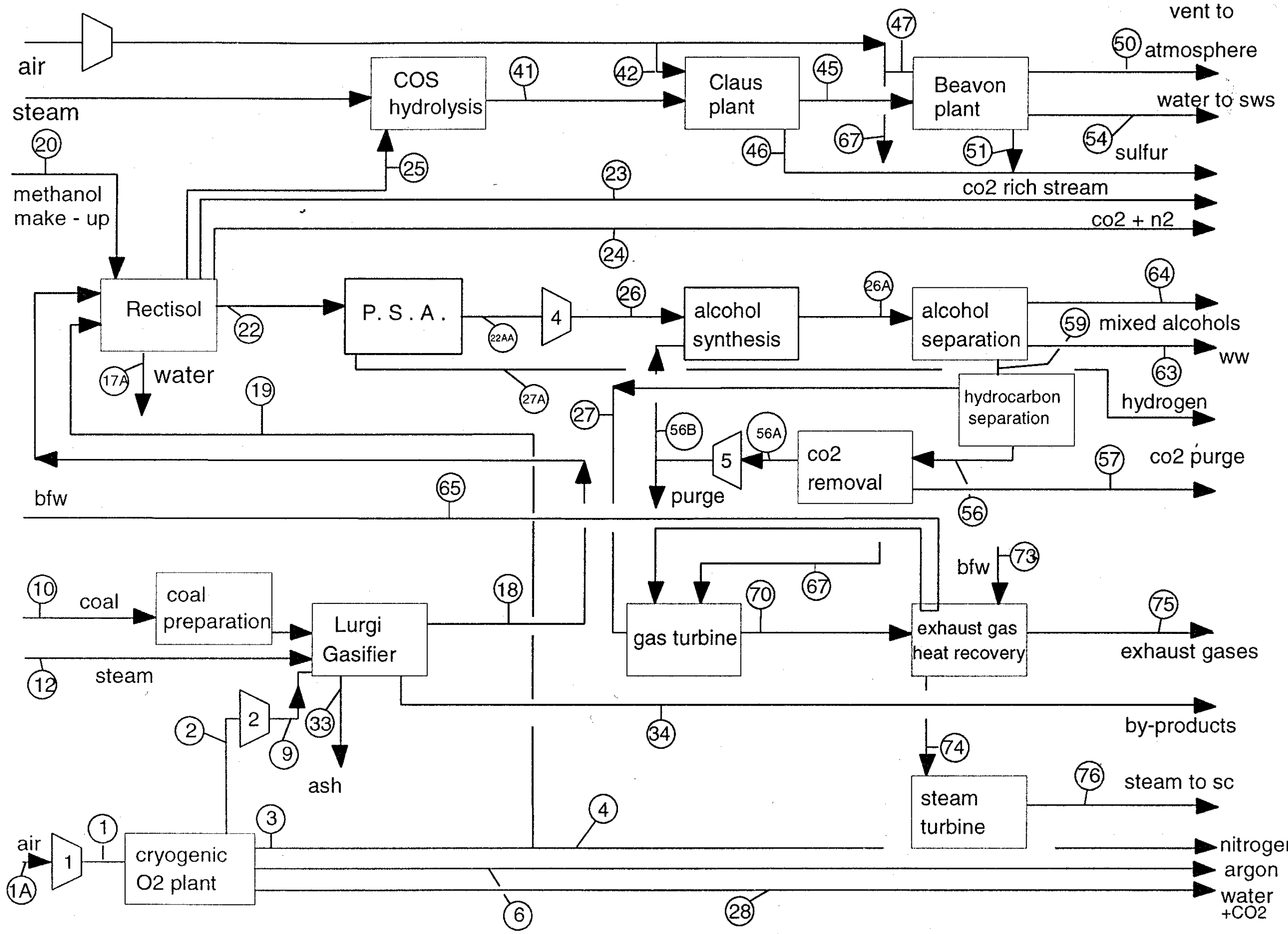

Table B.1 Case 2 Flow Table

	001	001A	002	003	004	006	009	010
Ar	372.8	372.8				372.8		
C								34157.0
CH3OH								
C2H5OH								
C3H7OH								
C4H9OH								
C5H11OH								
CO								
CO2	12.9	12.9						
COS								
CaCO3								
H2								13957.8
H2O	963.5	963.5						393.3
H2S								
N2	31216.9	31216.9		31216.9	29708.1			302.1
NH3								
O2	8392.5	8392.5	8392.5				8392.5	2690.0
S								459.0
Al2O3								639.5
C3H6O2								
C4H8O2								
CH4								
C2H6								
kmol/hr	40958.6	40958.6	8392.5	31216.9	29708.1	372.8	8392.5	52598.7
kg/hr	1175455.8	1175455.8	268559.6	874073.8	831826.2	14911.0	268559.6	619338.9
Temp.(C)	25.0	25.0	25.0	25.0	25.0	25.0	140.0	25.0
Press.(KPA)	500.0	101.3	500.0	500.0	500.0	500.0	2800.0	101.3

Table B.1 Case 2 Flow Table (cont'd)

	012	017A	018	019	020	022	022AA	023
Ar								
C								
CH3OH					5.8			
C2H5OH								
C3H7OH								
C4H9OH								
C5H11OH								
CO			11,261.9			11261.9	11261.9	
CO2			12,658.6					7595.2
COS								
CaCO3								
H2			18,922.4			18922.4	12674.3	
H2O	37985.6	23964.8	23,964.8					
H2S			459.0					
N2			299.9	1508.8				
NH3			4.4					
O2								
S								
Al2O3								
C3H6O2								
C4H8O2								
CH4			3,787.8			3787.8	3787.8	
C2H6			469.4			469.4	469.4	
kmol/hr	37985.6	23964.8	71828.3	1508.8	5.9	34441.6	28193.5	7595.2
kg/hr	683741.0	431366.8	1440290.1	42247.6	185.0	427866.1	415369.9	334187.4
Temp.(C)	400.0	25.0	25.0	25.0	25.0	25.0	25.0	25.0
Press.(KPA)	3900.0	2800.0	2800.0	500.0	101.3	2800.0	200.0	2800.0

Table B.1 Case 2 Flow Table (cont'd)

	024	025	026	026A	027	027A	028	033	034
Ar									
C								944.6	4,565.2
CH3OH		5.8		628.3					
C2H5OH				655.4					
C3H7OH				145.0					
C4H9OH				36.5					
C5H11OH				15.2					
CO			11261.9	7001.9	6594.0				
CO2	4430.5	632.9		1493.7	1406.7		12.9		
COS									
CaCO3									
H2			12674.3	7880.0	7421.0	6248.1			
H2O				140.1			963.5		
H2S		459.0							
N2	1809.8								
NH3		4.4							
O2									
S									
Al2O3								639.5	
C3H6O2				18.0					
C4H8O2				11.9					
CH4			3787.8	4490.5	4228.9				
C2H6			469.4	517.1	486.9				
kmol/hr	6239.3	1102.1	28193.5	23033.7	20137.5	6248.1	976.4	1584.2	4565.2
kg/hr	245588.7	43713.7	415369.9	432797.5	343640.2	12496.2	17911.5	76567.4	54781.9
Temp.(C)	25.0	25.0	240.0	310.0	25.0	25.0	25.0	25.0	25.0
Press.(KPA)	2800.0	2800.0	14000.0	12666.0	12666.0	2800.0	500.0	101.3	101.3

Table B.1 Case 2 Flow Table (cont'd)

	042	045	046	047	050	051	054	056	056A
Ar									
C									
CH3OH		5.8			5.8				
C2H5OH									
C3H7OH									
C4H9OH									
C5H11OH									
CO								407.9	407.9
CO2		632.9			632.9			87.0	
COS									
CaCO3									
H2								459.0	459.0
H2O		436.0					459.0		
H2S		22.9							
N2	820.1	820.1		109.6	929.7				
NH3		4.4			4.4				
O2	218.0			29.1	17.7				
S			436.0			22.9			
Al2O3									
C3H6O2									
C4H8O2									
CH4								261.6	261.6
C2H6								30.1	30.1
kmol/hr	1038.2	1922.2	436.0	138.7	1590.5	22.9	459.0	1245.6	1158.6
kg/hr	29940.6	59701.4	13952.9	4000.3	54705.8	734.4	8261.6	21256.2	17427.6
Temp.(C)	25.0	200.0	125.0	25.0	100.0	125.0	50.0	25.0	25.0
Press.(KPA)	101.3	101.3	101.3	101.3	101.3	101.3	101.3	12666.0	12666.0

Table B.1 Case 2 Flow Table (cont'd)

	056B	057	059	063	064	065	067	070
Ar								
C								
CH3OH					628.3			
C2H5OH					655.4			
C3H7OH					145.0			
C4H9OH					36.5			
C5H11OH					15.2			
CO	407.9		7001.9					
CO2		87.0	1493.7					13203.5
COS								
CaCO3								
H2	459.0		7880.0					
H2O				140.1		9633.9		26973.6
H2S								
N2							203424.9	203424.9
NH3								
O2							54075.0	36905.4
S								
Al2O3								
C3H6O2				18.0				
C4H8O2				11.9				
CH4	261.6		4490.5					
C2H6	30.1		517.1					
kmol/hr	1158.6	87.0	21383.2	170.1	1480.4	9633.9	257499.8	280507.3
kg/hr	17427.6	3828.6	364896.5	4905.1	62996.0	173411.0	7426295.3	7943346.5
Temp. (C)	45.0	25.0	25.0	25.0	25.0	25.0	25.0	590.0
Press. (KPA)	14000.0	12666.0	12666.0	12666.0	12666.0	10000.0	101.3	101.3

Table B.1 Case 2 Flow Table (cont'd)

	073	074	075	076	077	078
Ar						
C						
CH3OH						
C2H5OH						
C3H7OH						
C4H9OH						
C5H11OH						
CO						
CO2			13203.5			
COS						
CaCO3						
H2						
H2O	43501.6	43501.6	26973.6	43501.6	43501.6	43501.6
H2S						
N2			203424.9			
NH3						
O2			36905.4			
S						
Al2O3						
C3H6O2						
C4H8O2						
CH4						
C2H6						
kmol/hr	43501.6	43501.6	280507.3	43501.6	43501.6	43501.6
kg/hr	783028.2	783028.2	7943346.5	783028.2	783028.2	783028.2
Temp. (C)	25.0	535.0	200.0	40.0	380.0	535.0
Press. (KPA)	10000.0	10000.0	101.3	7.4	3000.0	3000.0

Table 8.2 Case 2 Energy Analysis

ELECTRICITY		
Plant	Electricity Used (MW)	Electricity Produced (MW)
Coal Preparation Plant	10.1	0.0
Cryogenic Oxygen Plant	11.2	0.0
Rectisol Plant	14.5	0.0
Lurgi Gasifier	3.1	0.0
Claus Plant	0.6	0.0
Gas Turbine	0.0	801.1
Steam Turbine	0.0	264.0
Compressor 1	81.4	0.0
Compressor 2	16.0	0.0
Compressor 4	146.0	0.0
Compressor 5	0.2	0.0
Total	285.1	1065.1