TECHNOLOGY ASSESSMENT GUIDE

NO. 6c

BGC LURGI GASIFICATION PROCESS

CHAPTER ONE: EXECUTIVE SUMMARY

1.1 OVERALL PROSPECTS FOR THE TECHNOLOGY

In 1955, the British Gas Council, in collaboration with the Ministry of Power, purchased an experimental three foot diameter slagging gasifier which the German Lurgi company had been developing. Testing was carried out until 1964, followed by an eleven year hiatus. In 1975, BGC resumed testing under sponsorship of the Conoco Development Company and other U.S. sponsors.

For the high-Btu gasification plant discussed here, the technology employed in the remainder of the plant is strictly conventional, with the gasifier being the only critical item.

In comparison to the non-slagging version, the BGC-slagging Lurgi gasifier has a much higher volumetric gas throughput (approximately four times as great), consumes much less steam, produces a gas of higher heating value, and displays higher thermal efficiency.

On the negative side, oxygen consumption is slightly greater, as is tar yield. For most applications, these draw-backs are minor in comparison to the advantages offered by this approach over the conventional Lurgi design.

The lack of commercial operating experience is a highly important factor weighing against the BGC reactor in choosing a gasifier as the basis for a new plant. However, considerable experience has been gained through the years of operating history at the pilot plant scale, and many observers feel that this system is now ready for full-scale implementation. For applications involving synthesis gas generation for methane or chemicals manufacture, combined cycle gasification, and industrial fuel gas generation, the BGC gasifier is fully competitive on technical and economic grounds with other advanced generation gasifiers.

1.2 ENGINEERING ASPECTS

The current BGC experimental gasifier (located at Westfield, Scotland) is similar in design to the Lurgi dry ash gasifier, but is only 6 ft (rather than 9 ft) in diameter. Oxygen and steam injection are accomplished through tuyeres located in the gasifier wall toward the bottom of the bed. A water jacket encircles the vessel, and refractory lining is used inside the steel vessel wall. Slag is intermittently withdrawn from a slag tap hole at the bottom of the bed, below which it is quenched with water in an unlined vessel. Slag and quench water removal takes place through a slag lockhopper located below the quench vessel. This slag removal and quench design is one of the major differences between this and the rotating grate dry ash removal of the conventional Lurgi system. Due to the extremes of temperature in the slagging gasifier and the abrasive nature of the ash, the slag tap hole will experience a high erosion rate. The refractory lined walls of the gasifier will also require frequent replacement, due to the fragility of the insulating material.

Coal is introduced at the top of the reactor through a lockhopper and distributor arrangement, similar to the nonslagging version. In choosing a coal feedstock for either the slagging or nonslagging gasifier, several considerations apply:

- Coal should range in size from 1/8 to 1-1/2 inch
- Moisture level should be below 35 percent
- Noncaking coals should be used unless the feed is pretreated or there is a mechanical stirrer in the gasifier
- Up to 10 percent coal fines (<1/8") may be used if they are injected with the steam at the bottom of the bed.

Due to its fixed bed design, the slagging version is similar to the nonslagging gasifier in the extent of production of tars, oils, phenols, naphthas, ammonia and hydrocarbons, and will therefore require extensive wastewater treatment. The volume of wastewater will be significantly lower in the case of the slagging Lurgi, because of its drastically lower consumption of steam. This will effectively increase the concentration of organics in the wastewater which is produced, which will require a different treatment system.

1.3 CURRENT COSTS

The total capital requirement for this 91.25×10^{12} Btu/year plant is $2.2 billion, which is dominated by a plant investment of $1.4 billion and interest during construction of $555 million. Start-up costs and working capital each contribute approximately $85 million.

Annual operating and maintenance costs (at a 90% plant capacity factor) total $120 million. By-product credits for sulfur, ammonia, tar-oil, phenols and naphtha total $45 million annually, offsetting the total operating costs to $75 million net. These costs are exclusive of coal costs, and hence do not reflect coal feedstock expenditures or creditks for coal fines sold.

Taken together with a 20 percent capital charge, these operating costs result in a product cost of $5.35/10^6 Btu, which is exclusive of coal costs. This compares favorably with the Exxon CCG process, and is within the same range of uncertainty as the costs for the IGT HYGAS process.

As a reflection on the superiority of the slagging version, the production costs for this case are roughly half of that for the nonslagging Lurgi-ANG facility (see TAG No. 5). However, there is a greater degree in the uncertainty associated with these costs (for the slagging reactor) due to the fact that the ANG plant is closer to commercialization.

1.4 RESEARCH AND DEVELOPMENT DIRECTIONS

The slagging Lurgi design represents overall a significant improvement on the original nonslagging design. The major objective at this stage of development is to demonstrate reliable and efficient gasifier performance over long periods of time with different coals. Beyond this, it may be possible to improve the lifetime of refractory liners by the use of new materials of construction. The slag tap hole will also be an area of concern. Useful recovery of heat from the slag quench operation would aid overall process efficiency, and should be an early goal. Additional improvements in efficiency may result from a further reduction in steam usage. The net tar yield is higher than in the nonslagging version, due to the radiative transfer of heat to the incoming coal from the high temperature zone. Injection of high temperature steam and possibly oxygen in the devolatilization zone may gasify some of these organics to more useful gaseous compounds rather than by-product liquids which foul process equipment and wastewater streams.

CHAPTER TWO: ENGINEERING SPECIFICATIONS

2.1 GENERAL DESCRIPTION OF THE TECHNOLOGY

In the BCG-Lurgi gasification process, coal sized to approximately 1" is introduced through a lockhopper and distributor arrangement into a pressurized fixed bed gasification vessel. The coal reacts with steam and oxygen in the gasifier to produce raw synthesis gas, which is discharged through the top of the vessel. Molten slag is withdrawn from the bottom of the gasifier and is quenched with water.

The synthesis gas is purified, cooled, desulfurized, and methanated to produce a high-Btu pipeline-quality gas. The cooling of the raw gas produces a gas liquor condensate from which oils, phenols, and ammonia are recovered.

In 1953, Lurgi Kohle und Mineraloeltechnik GmbH began experimenting with a slagging fixed bed gasifier. In 1955, an experimental gasifier of this type was purchased by the British Gas Council in collaboration with the Ministry of Power. Development tests were conducted in England until 1964, when the North Sea gas discovery was made. The British Gas Corporation resumed slagging gasification development in 1974 through funding by the Conoco Coal Development Company and other U.S. sponsors.

The commercial plant assessed herein is based on commercially-proven processes and technology except for the gasification and methanation steps. The gasification process is based on technology held by British Gas Corporation and Lurgi GmbH. The methanation technology is held by Conoco Methanation Company and was demonstrated on a semicommercial scale in 1973-74 at Westfield, Scotland.[2-1]

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2.2 PROCESS FLOW, ENERGY, AND MATERIAL BALANCES

Relevant plant area numbers for the BGC-Lurgi process are listed in Table 2-1. The interaction of these units in overall plant operation is illustrated by the conceptualized process flow diagram of Figure 2-1, while the compositions of process streams in the plant are listed in Table 2-2.

As shown in Figure 2-1, coal as received is stockpiled, ground, and classified by size prior to delivery to the gasifier. Steam, which is generated by a tail gas combustion unit, and oxygen, produced by an onsite air separator, are reacted with the coal in the gasifier at pressures of 5 to 26 atm. Molten slag is removed from the bottom of the gasifier and quenched prior to disposal. Raw gas leaves the top of the gasifier and is piped to a shift conversion unit, which raises the H_2:CO ratio in the gas. The converted gas is sent to a quenching and cooling unit, which produces an oily liquor condensate and a cool gas stream. Oil product, phenols, and ammonia are separated and recovered from the oily liquor, while the gas stream is sent to a gas cleaning unit which removes naphtha product, acid gas, and CO_2. Elemental sulfur is extracted from the acid gas stream in a recovery unit and conveyed to storage. The desulfurized synthesis gas, composed primarily of H_2 (70% $^v/_v$) and CO (21% $^v/_v$), is methanated in a unit which effects the reaction

$$CO + 3H_2 \longrightarrow CH_4 + H_2O.$$

The resulting product, composed of 94 percent methane, is compressed to produce a pipeline-quality gas.

A plant material and energy balance analysis for the BGC-Lurgi complex is presented in Table 2-3. An overall plant efficiency of approximately 60 percent is expected based on a pipeline gas, naphtha, oil, phenol, and ammonia product output of 250 billion Btu/day.

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Table 2-1

Relevant BGC-Lurgi Plant Area Numbers

100 COAL STORAGE AND HANDLING

- 110 Coal Storage
- 120 Coal Handling and Storage

200 COAL PREPARATION

- 210 Coal Grinding
- 250 Size Classification

300 GASIFICATION

- 310 Gasification
- 320 Slag Quench

1200 RAW GAS COOLING

- 1220 Gas Quenching and Cooling

1300 ACID GAS REMOVAL AND GAS CLEANING

- 1310 H_2S and CO_2 Removal
- 1320 Ammonia Recovery
- 1330 Tar and Oil Separation
- 1340 Phenol Recovery

1400 SULFUR RECOVERY AND TAIL GAS TREATING

- 1410 Sulfur Recovery

1600 PRODUCT GAS COMPRESSION

1700 SHIFT CONVERSION

1800 METHANATION

1900 AIR SEPARATION

2000 UTILITIES AND SUPPORT SYSTEMS

- 2010 Steam Generation and Power Recovery
- 2020 Wastewater Treatment
- 2050 Water Treatment
- 2060 Flue Gas Desulfurization

2100 OFFSITES AND MISCELLANEOUS

- 2140 Cooling Towers

Figure 2-1

HGC-Lurgi Conceptualized Process Flow Diagram

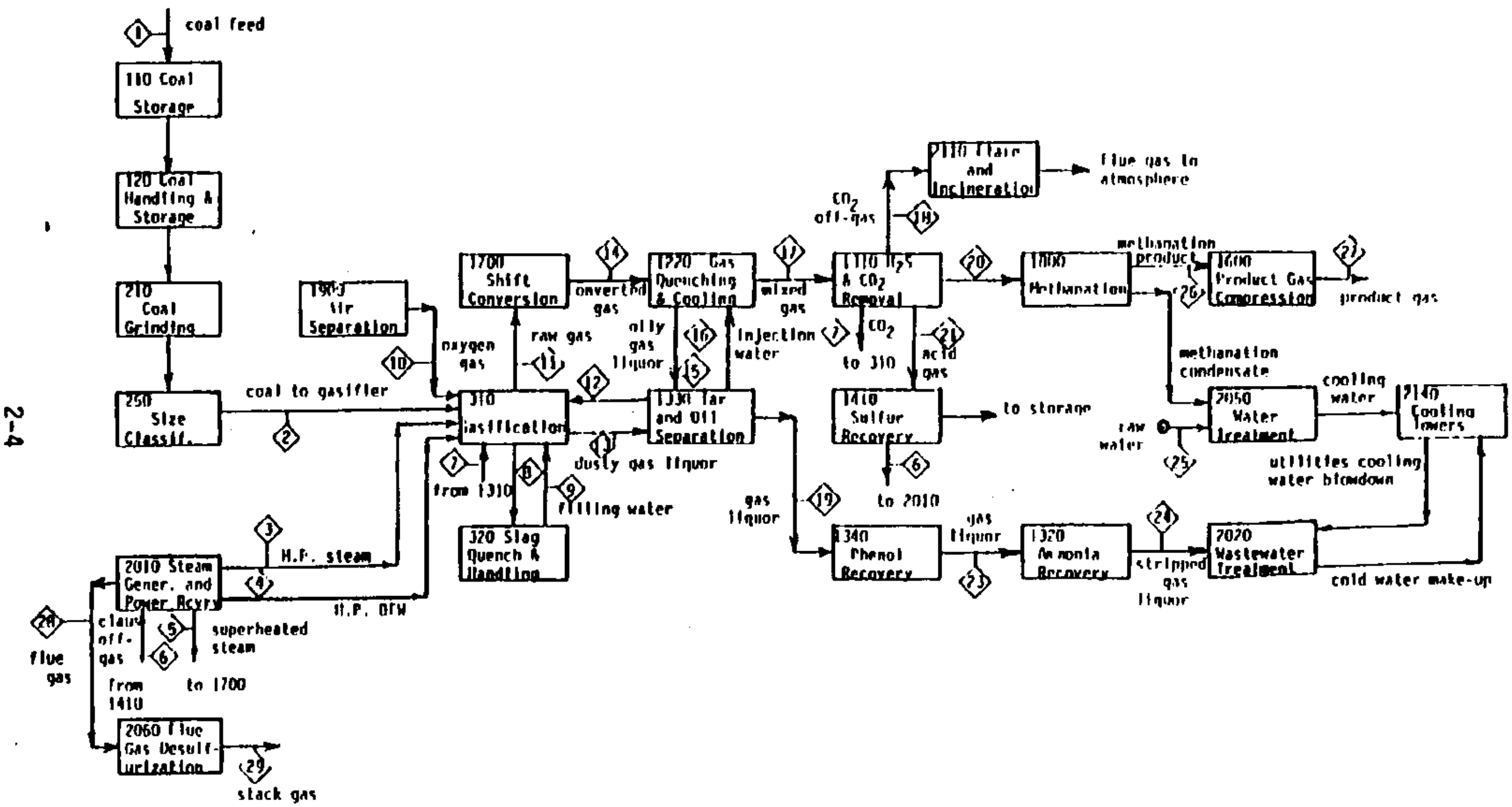

Table 2-2

BGC-Lurgi Detailed Material Balance for Process Streams

Stream No.	1		2		3		4		5		6		7		8		9	
Description	Coal feed		Coal to gasifier		H.P. steam		H.P. BFW		Superheated steam		Claus off-gas		Carbon dioxide		Slag and water		Filling water	
Temperature, °F			77		750		250		750		320		158		158		158	
Pressure, PSIG					550		795		550		0.2		500		12			
Flow, klb/hr mole %	klb/hr		klb/hr		klb/hr		klb/hr		klb/hr		klb/hr	mole %	klb/hr	mole %	klb/hr		klb/hr	
CO											3.392	2.39	0.264	0.19				
CO_2											119.004	54.93	215.270	98.43				
H_2											0.124	1.25	0.018	0.18				
CH_4											0.082	0.10	0.471	0.59				
C_nH_m											1.333	0.50	0.687	0.59				
H_2S											1.086	0.65	0.005	(>0.01)				
S											0.930	0.14	0.046	0.02				
N_2											54.767	39.72						
O_2																		
SO_2											1.022	0.32						
Other																		
Total Dry Gas											181.650		216.761					
Water											38.316							
Total Wet Gas																		
Ammonia																		
Phenol																		
Naphtha																		
Solvent																		
Other Compounds																		
Total Stream	2413.427		1475.720		461.75		248.168		914.894		219.966		216.761		497.592		375.000	

Table 2-2

BGC-Lurgi Detailed Material Balance for Process Streams (cont'd)

Stream No.	10		11		12		13		14		15		16		17		18	
Description	Oxygen gas		Raw gas		Injection water		Dusty gas liquor		Converted gas		Oily gas liquor		Injection water		Mixed gas		Carbon dioxide off-gas	
Temperature, °F	275		331				356		878		226				95		68	
Pressure, PSIG	500		421				250		379		250				355		1.2	
Flow, klb/hr, mole %	klb/hr	mole %	klb/hr	mole %	klb/hr		klb/hr	mole %	klb/hr	mole %	klb/hr	mole %	klb/hr	mole %	klb/hr	mole %	klb/hr	mole %
CO			1474.247	58.52			0.327	11.32	454.815	12.86	0.130	0.43			492.970	13.81	1.788	0.19
CO_2			254.985	6.44			1.844	40.52	1859.864	33.48	41.647	88.23			1824.830	32.52	1456.739	98.63
H_2			46.574	25.69			0.012	5.80	119.098	46.81	0.028	1.30			120.280	46.80	0.120	0.18
CH_4			87.892	6.09			0.022	1.35	93.590	4.62	0.058	0.34			98.814	4.69	3.185	0.59
C_nH_m			16.265	0.50					11.409	0.25					11.831	0.26	4.651	0.39
H_2S			59.105	1.93			1.455	41.01	62.816	1.45	3.548	9.70			60.173	1.39		
S			6.475	0.12					1.271	0.02					1.439	0.02	0.310	0.02
N_2	11.379	2.00	17.913	0.71					17.913	0.51					18.378	0.51	0.030	(>0.01)
O_2	636.904	98.00																
SO_2																		
Other																		
Total Dry Gas	648.288		1963.456				3.650		1620.146		45.411				2625.723			
Water			531.734				828.218		725.766		974.421				5.154			
Total Wet Gas			2495.190				831.868		1345.912		1019.832				2630.877		1466.823	
Ammonia																		
Phenol																		
Naphtha																		
Solvent																		
Other Compounds			39.042				115.294		36.706		22.769				14.949			
Total Stream	648.288		2534.232		737.503		947.862		1382.618		1042.601		120.000		2645.826		1466.823	

Table 2-2

BGC-Lurgi Detailed Material Balance for Process Streams (cont'd)

Stream No.	19		20		21		22		23		24		25		26		27	
Description	Gas liquor		Synthesis gas		Acid gas		Sulfur product		Gas liquor		Stripped gas liquor		Raw water		Methanation product		Product gas	
Temperature, °F	95		73		68				96						230		102	
Pressure, PSIG	39		280		10				157						205		1032	
Flow, klb/hr, mole %	klb/hr	mole %	klb/hr	mole %	klb/hr	mole %	klb/hr		klb/hr	mole %	klb/hr		klb/hr		klb/hr	mole %	klb/hr	mole %
CO			490.918	20.52											0.012	(>0.01)	0.012	(>0.01)
CO_2	6.116	75.08	77.965	2.07	74.864	48.15			8.979	73.97					9.321	0.80	9.296	0.80
H_2			120.142	69.77											1.314	2.45	1.311	2.45
CH_4			92.158	6.72											402.426	94.29	401.349	94.29
C_nH_m			3.825	0.15	2.666	1.38												
H_2S	1.572	24.92			60.173	49.98			2.447	26.03								
S					1.044	0.49												
N_2			18.345	0.77											18.345	2.46	18.296	2.46
O_2																		
SO_2																		
Other																		
Total Dry Gas	7.688		803.353		138.747				11.426						481.102		430.325	
Water	1016.746								1044.189									
Total Wet Gas	1024.434								1055.615									
Ammonia									6.089									
Phenol									.633									
Naphtha																		
Solvent									10.550									
Other Compounds	15.513								4.685									
Total Stream	1039.947		803.353		138.747		75.566		1077.572		904.557		5102.022		481.102		430.325	

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Table 2-2

BGC-Lurgi Detailed Material Balance for Process Streams (cont'd)

Stream No.	28		29	
Description	Flue gas		Stack gas	
Temperature, °F				
Pressure, PSIG				
Flow, klb/hr, mole %	klb/hr	mole %	klb/hr	mole %
CO				
CO_2	911.550	16.65	911.550	16.73
H_2				
CH_4				
C_nH_m				
H_2S				
S				
N_2	2763.397	79.32	276.397	79.68
O_2	163.508	3.43	136.342	3.44
SO_2	38.053	0.48	2.356	0.03
Other	0.598	0.12	0.598	0.12
Total Dry Gas	3850.106		3814.243	
Water	263.702		407.878	
Total Wet Gas	4113.808		4222.121	
Ammonia				
Phenol				
Naphtha				
Solvent				
Other Compounds	1.365			
Total Stream	4117.173		4222.121	

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Table 2-3

Overall Material and Energy Balance

	Mass Flow Rate klb/Hr	Gross Heating Value MM Btu/Hr
Input		
Coal to Gasification	1,406.566	14,332.50[a]
Coal to Boilers	374.000	3,810.95[a]
Excess Coal Fines	563.707	(b)
Total Input	2,344.273	18,143.45
Products		
Pipeline Gas	430.325	9,666.79[c]
Naphtha	14.988	265.64
Oil	21.873	373.24
Crude Phenols	5.712	73.29
Anhydrous Ammonia	4,010	38.77
Sulfur	76.566	307.40
Sodium Sulfate Purge	3.835	—
Coal Fines	563.707	(b)
Total Products	1,121.016	10,725.13
Process Thermal Efficiency		59.1%

$$\text{Overall Plant Efficiency} = \frac{10,725}{18,143} = 59.1\%$$

[a]Illinois No. 6 Seam, HV = 13650 Btu/lb (DAF),
HV = 10190 Btu/lb as received

[b]The excess coal fines which are sold have not been included in the plant energy balance to avoid distorting the plant efficiencies. These coal fines are an additional feed and product of 5,744.01 MM Btu/Hr.

[c]Product gas heating value = 22463 Btu/lb.

Source: Reference 2-2

2.3 PLANT SITING AND SIZING ISSUES AND CONSTRAINTS

The plant would be located near the coal mine within easy delivery of "run-of-mine" coal by overland conveyor. Approximately 2,560 acres are required for the plant. Coal preparation and storage, processing units, utility plants, equilization basins, product storage, and general facilities will occupy 480 acres. An onsite air separation facility capable of producing 648,000 lbs/hr of 98 percent pure oxygen will be required. An estimated 1,800 acres are required for solids waste disposal for a 20-year period.

Access to the plant from highways, primary roads, and railroads is a major consideration. A transportation study will be necessary.

2.4 RAW MATERIALS AND SUPPORT SYSTEMS REQUIREMENTS

2.4.1 Coal Quantities and Quality

The feedstock chosen by the Continental Oil Company for their studies is an Illinois No. 6 Seam coal. The Illinois No. 6 coal was selected for the commercial plant design because it is representative of large coal reserves found in Illinois, Indiana, and Kentucky. The throughput to the gasifier is 16,879 tons per day. An additional 4,488 tons per day are consumed as fuel for on-site steam/power generation. The quality of this bituminous feedstock is illustrated in Table 2-4.

2.4.2 Catalysts and Other Required Materials

The required catalysts and chemicals for the various facilities within the plant are presented in Table 2-5. Included in this table are such major facilities as the water treatment and steam generation system, the cooling water system, and the wastewater treatment system.

2.4.3 Water Requirements

Table 2-5 presents a summary of the water requirements for the individual facilities within the commercial plant. The requirements for the H_2S and CO_2 removal unit, product gas compression and drying system, the water treatment and steam generation facility, and other units are shown in this table.

Table 2-4

Properties Of Coal For Plant Design

Type	Illinois No. 6
Proximate Analysis	Weight Percent
Moisture	12.08
Ash	13.27
Volatiles	30.80
Fixed Carbon	43.85
	100.00
Ultimate Analysis (DAF Basis)	
Carbon	76.55
Hydrogen	5.26
Oxygen	10.92
Nitrogen	1.11
Sulfur	5.95
Chlorine	0.21
	100.00
Coal Heating Value (DAF Basis)	13,650 Btu/Lb
Ash Fusion Characteristics (Reducing)	°F
Softening Point	1,911
Melting Point	1,980
Flow Point	2,575

Source: Reference 2-2

Table 2-5

Catalysts and Chemicals

Facility	Chemical	Rate
Coal and Flux Handling and Preparation	Dust Separation Liquid	4200 gal/year
Rectisol	Methanol	700 lb/hr
	Sodium Hydroxide (20% by wt.)	500 "
Product Gas Compression and Drying	Triethylene Glycol	9.8 "
Phenol Extraction	L.P. Nitrogen	5.0 MSCF/hr
	Solvent (Isopropyl Ether)	6.0 lb/hr
Water Treatment and Steam Generation	Alum (50 wt%)	125.0 "
	Polymer (Dry Polymer)	10.0 "
	Hydrated Lime (74 wt%)	600.0 "
	Hypochlorite (as Cl_2)	4.0 "
	Sulfuric Acid (66° Beaume)	1000.0 "
	Sodium Hydroxide (50 wt%)	800.0 "
	Tri-Sodium Phosphate	22.0 "
	Morpholine (40 wt%)	20.0 "
	Hydrazine (35 wt%)	0.5 "
Cooling Water System	Chlorine (Gas)	12.5 "
	Dispersant (Active Ingredients)	70.8 "
	Chrome Inhibitor (Contains CrO_4)	41.7 "
	Zinc Inhibitor (Contains Zn)	12.5 "
	Sulfuric Acid (66° Beaume)	50.0 "
	Sodium Hydroxide (55% by wt.)	83.3 "
Wastewater Treatment	Lime (as CaO)	275.0 "
	Polymer	10.0 "
	Aqueous Ammonia (20% by wt.)	As required
	Phosphoric Acid (54% by wt.)	70.0 lb/hr
	Fixation Chemicals	1800.0 "
	Activated Carbon	270.0 "

Source: Reference 2-1

Table 2-6

Water Requirements

Plant Area No.	Facility	Type of Water	Rate	Battery Limit Conditions
1310	H_2S and CO_2 Removal	Cooling Water	9000 GPM	65 psig, 87°F
		Medium Pressure Boiler Feed Water	66.0 KLB/HR	190 psig, 250°F
1600	Product Gas Compression and Drying	Cooling Water	19,940 GPM	100 psig, 87°F
2050, 2010	Water Treatment and Steam Generation	Cooling Water	23,400 GPM	65 psig, 87°F
2020	Wastewater Treatment	Cooling Water	31,710 GPM	65 psig, 87°F

Source: Reference 2-3

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2.5 EFFECT OF COAL TYPE

The Conoco Coal Development Company has performed studies with the British Gas/Lurgi Slagging Gasifier using three different kinds of coals. An Ohio No. 9 unwashed coal was fired, in order to investigate the gasifier operability with a high ash feed. Pittsburgh No. 8 coal was washed before it was tested. Frances coal from Scotland was the third feedstock evaluated. The Frances coal was also washed; however unlike the Pittsburgh coal, it is not a highly caking feed.

Eastern bituminous coals are relatively unreactive and have lower ash-fusion temperatures than many Western coals. The elevated slagger temperature is ideally suited to deal with these properties since reactivity has little influence at the high temperatures in the slagger. In addition, withdrawal of ash as a fluid eliminates concern about partial fusion and sticking to a bottom grate.

Experience with the non-caking, washed Frances coal was satisfactory throughout the Conoco test runs. Several runs were required with the American coals in order to determine their operating characteristics. The Ohio No. 9 feed varied considerably in composition as fed to the gasifier. There are no tabulated results of the gasifier performance for this feedstock, however those runs were successful.

The Pittsburgh Seam test gave no evidence of problems during its duration. These results were equivalent to those obtained with Frances coal, indicating that the facility is able to compensate for significant differences in coal caking properties as well as ash contents and compositions. Table 2-7 summarizes the inspections of the Frances and Pittsburgh coals; most critical to the operation of the gasifier are ash content, fluid temperature, and the coal free swelling index. Table 2-8 presents the corresponding gasification results. The Pittsburgh Seam results are equivalent to those obtained with the Frances coal, indicating that the facility is able to compensate for significant differences in coal caking properties as well as ash contents and compositions.

Table 2-7

Westfield Lurgi Performance Data

	Coal Inspections	
Shaft Cross Section	28.2	28.2
Coal	Pgh. No. 8	Frances
Size Consist	1-1/4" x 1/4"	1" x 5/8"
Free Swelling Index	8.0	0
Prox., Dry Basis		
% Voltatile	40.8	35.9
% Fixed Carbon	51.9	59.3
% Ash	7.3	4.8
Ash Fluid Point, °F	2680	2630
Ash Iron Content, % Fe_2O_3	19.1	13.2

Source: Reference 2-3

Table 2-8

Westfield Operating Results

Coal	Pittsburgh No. 8	Frances
Operating Conditions		
Pressure, psig	350	350
Rates, Ton/Ton MAF Coal		
Oxygen	0.56	0.54
Steam	0.41	0.41
Fuel Rate Lb (MAF)/Hr/Sq Ft	870	852
Yields, Per Ton MAF Coal		
Tar + Oils, Lb	NA	255
Liquor	NA	400
Raw Gas, MSCF	72	70
Gas Gross HV, Btu/CF	356	352
SNG Equivalent, MSCF	20.3	20.4
Run Duration, Hrs	88	524
Post-Run Inspection	Good	Good

Source: Reference 2-3

2.6 AIR POLLUTION CONTROL TECHNOLOGY

The flue gas desulfurization unit (plant area no. 2060) utilizes the Wellman-Lord sulfur dioxide recovery process, a closed loop regenerative process. In this process a sodium sulfite solution absorbs sulfur dioxide from the gas stream. The principle reaction is as follows:

$SO_3 + Na_2SO_3 + H_2O = 2NaHSO_3$ (reference 2-3).

The circulating solution is regenerated by evaporation. The primary reaction occurring in the evaporators is the following:

$2NaHSO_3 = Na_2SO_3$ (crystals) $+ SO_2 + H_2O$ (reference 2-3).

The boilers also incinerate sulfur plant tail gas, therefore the Wellman-Lord process eliminates the need for a sulfur plant tail gas unit. The unit will process the total flue gas from the boilers while the boilers operate at maximum capacity. The flue gas desulfurization unit reduces the sulfur dioxide to a concentration of 251 ppm by volume.

The sulfur recovery unit (plant area nc. 1410) converts sulfur compounds, mainly hydrogen sulfide and sulfur dioxide, to elemental sulfur. The conversion is accomplished by the following reaction:

$2H_2S + SO_2 \rightleftarrows 3S + 2H_2O$.

Sulfur compounds are kept from polluting the atmosphere by recovering the sulfur in its elemental form.

The system is designed to recover 97 percent by weight of the sulfur in the feed. The process streams that constitute this sulfur are:

Feed Streams	Tons/Day
Acid Gas (H_2S and CO_2 Removal)	686.0
Phenol Extraction Acid Gas	17.5
Boiler Flue Gas SO_2	203.1
Gas Liquor Separation Expansion Gas	41.6
TOTAL	948.2

The Flare System (plant area no. 2110) is designed to provide safe burning of combustible vapors produced during programmed start-up and shutdown, or during upset conditions. Unit upsets include emergencies such as power failure, cooling water failure, unit depressurization, instrument air failure, fire, and blocked outlets.

Two gaseous waste streams, the off-gas from H_2S and CO_2 removal and lock gas from gasification are processed in the incineration unit. The incineration of the carbon dioxide rich streams, in the presence of excess air, is accomplished at a minimum temperature of 1500°F and a residence time of 0.8 seconds.

Systems for preheating the off-gas and a portion of the combustion air reduce the need for supplemental fuel. Both of these systems make use of incinerator flue gas as the heating medium. Heat is also recovered from the flue gas, through the waste heat boiler, before entering the preheaters. After heat exchange with the air and off-gas, the flue gas is released to the atmosphere by means of a 100-foot tall stack, in order to ensure satisfactory dispersion. The only pollutant emitted from the stack is sulfur dioxide, at a concentration of 136 ppm by volume.

2.7 WATER POLLUTION CONTROL TECHNOLOGY

2.7.1 Ability of the Existing Technology to Meet Regulations

The commercial plant is designed to utilize commercially available wastewater treatment processes. The wastewater treatment system permits the recycle and reuse of treated water to meet the "zero discharge" requirement stipulated by the DOE. All plant liquid effluent streams are collected and treated by the system. Primary, secondary, and tertiary wastewater treatment methods are incorporated into the system.

2.7.2 Water Recycling Systems

The water treatment process utilizes conventional equipment to process water for use in the following systems.

1. Firewater (4000 gpm max)
2. Service Water (300 gpm)
3. Potable Water (300 gpm)
4. Cooling Water (8640 gpm)
5. Boiler Feed Water (2935 gpm)

The main features of the water treatment system are raw water clarification, filtration, demineralization, condensate polishing, and boiler feed water deaeration.

2.8 SOLID WASTE HANDLING

2.8.1 Disposal Requirements

As noted in section 2.7, Water Pollution Control Technology, the commercial plant has been designed for zero discharge of wastewater. Only gaseous emissions to the atmosphere and solid waste disposal will have an effect on the plant's surrounding environment. The plant design incorporates the best available technology in order to minimize any adverse effects.

The following table presents those sections where solid wastes are produced.

Plant Area No.	Section	Source	LB/HR
320	Slag Quench and Handling	Slag	251451
2050	Water Treatment	Coal Ash	46265
		Vacuum Filter Dewatering	7765
		Sulfate Purge	3835
2020	Wastewater Treatment	Sludge and Coal Ash	65275

2.8.2 Leachate Problems

Solid wastes are to be buried on the plant site in landfill operations. About 90 acres per year are uncovered for these operations. The top soil will be removed and temporarily stored, until the landfill operations have been completed. At that time the top soil will be replaced and the land replanted. Best Available Technologies will be applied to minimize leaching and groundwater contamination.

2.9 OSHA ISSUES

Workers may be exposed to coal dust and to noise while handling and preparing the coal for gasification. Coal dust, which can cause black lung disease, can be controlled by wetting the coal pile. A risk of fire exists where dried, ground coal is stored.

Many of the liquid by-products of the slagging Lurgi process are likely to be carcinogenic. Coal tar contains polynuclear aromatic hydrocarbons, including benzopyrene, a strong carcinogen and benzene, which can cause leukemia. Phenols, a major by-product, are co-carcinogens, substances which are not themselves carcinogenic, but will promote multiplication of abnormal cells after initiation of the conversion of a normal cell to a latent tumor cell by an active carcinogen.[2-4] Therefore, protective clothing must be worn and strict personal hygiene must be maintained in areas where the liquid by-products are handled. Exposure risk will be high during equipment cleaning and maintenance.

2.10 PROCESS PERFORMANCE FACTORS

The commercial plant produces 10,725 MM Btu per hour of products from 14,332 MM Btu per hour of coal feed to the gasifiers. The overall plant thermal efficiency is 59.1 percent, while the process thermal efficiency is 74.8 percent. The overall thermal efficiency is the more significant figure. The energy required to drive the pumps and compressors accounts for this difference.

The overall plant efficiency could be improved through reference to recent experience on a pilot plant gasifier and through improving the design for the steam/power plant. The data generated from these recent studies indicate that the oxygen requirement for gasification is 10 percent less than that specified in the commercial plant design.

2.11 TECHNOLOGY STATUS AND DEVELOPMENT POTENTIAL

The program is in the first of three phases, Development and Engineering. BGC has demonstrated that the elevated slagger temperature is ideally suited for the highly unreactive Eastern bituminous coals with relatively low ash-fusion temperatures. In addition, a test program on weakly-caking British coals and Western U.S. coals was successfully completed in early 1977 by a consortium of 14 U.S. industrial firms.

The commercial plant design maximizes the use of commercially available processing technology to minimize the overall plant risk. Most of the process units have been operated commercially in large plants.

The gasification unit is the new process development area in the plant. The scale-up of the gasifier from the pilot plant to the demonstration plant size is a factor of 2.8 to 1, the scale-up of the gasifier from the demonstration plant to the commercial plant is 1.44 to 1. The gasifier is the least proven unit in the plant design.

The shift conversion unit processes more gas than previous shift convertors. Scale-up is not a problem because the unit consists of conventional heat exchangers and fixed-bed reactors. The only potential problem is the low hydrogen to carbon monoxide ratio of the feed, which could result in fouling. However, because this has been considered in the design of the first shift reactor, the risk is defined as low-to-moderate relative to the rest of the units in the plant.

The methanation system was demonstrated in Westfield, Scotland, where 2.5 MMSCF per day of product SNG were produced. The 30 to 1 scale-up factor, on a product basis, presents no problem because the equipment is all conventional, and the design configuration is such that the reactor scale-up factors are about 4 to 1. Because of the somewhat limited experience, the risk is considered low-to-moderate relative to the rest of the plant.

The Wellman-Lord flue gas desulfurization system is proven in many applications. The scale-up requirement is about 1.5 to 1, and the risk is low-to-moderate.

With the exception of the "zero discharge" equipment, the wastewater treatment area consists of commercial equipment. The design is conservative enough to compensate for any reasonable error in the assumptions made during the design. The risk factor in the conventional wastewater treatment area is low-to-moderate. The "zero discharge" criteria adds additional equipment to the process. The disposal of the solids by chemical fixation is the greatest unknown with a moderate-to-high risk factor.

2.12 REGIONAL FACTORS INFLUENCING ECONOMICS

2.12.1 Resource Constraints

The restriction in the use of coal feedstocks constrains to some extent the range of acceptable coals for the plant, and because of this makes the plant somewhat more vulnerable to price escalation or supply interruptions. However, the use of water recycling in the plant provides some degree of freedom in locating the plant near a diversity of acceptable coal sources, even if these happen to be in arid regions or in areas where water supplies are restricted for other reasons.

2.12.2 Environmental Control Constraints

Regulations governing air and water quality could be quite significant in their effect on plant economics. Regulations governing solid waste may also be important but probably less so than those covering air and water quality. Due to the nature of the fixed bed gasifier (high organic concentrations in off-gas condensate), considerable attention to water quality and wastewater treatment will be required. In the likely event that a zero liquid discharge criterion is imposed, several options exist (including shifting the burden to air emissions) but all are associated with higher capital and/or operating costs. The exact impact on process economics can only be determined from a knowledge of the regulations applying to a specific site.

2.12.3 Siting Constraints

Site selection may be an important factor in determining project financial success by virtue of the impacts of resource supply and environmental control regulations. Ease of construction (and scheduling), taxation rates, proximity to barge, rail and pipeline gas networks, and labor cost and availability are all factors which will influence the impact of site selection on project economics.

References

2-1. Phase 1: The Pipeline Gas Demonstration Plant, Design and Evaluation of Commercial Plant, Volume 2. Process and Project Engineering Design, Continental Oil Company, Stamford, Connecticut, FE-2542-10 (Vol. 2).

2-2. Phase 1: The Pipeline Gas Demonstration Plant, Design and Evaluation of Commercial Plant, Volume 1. Executive Summary, Continental Oil Company, Stamford, Connecticut, FE-2542-10 (Vol. 1).

2-3. Sudbury, J.D., G.P. Curran, and J.A. Phinney. Gasification of Caking Coals in the BGC/Lurgi Slagging Gasifier -- A Status Report --, Conoco Coal Development Company, Research Division, presented at the Fifth Annual International Conference on COAL GASIFICATION, LIQUEFACTION, AND CONVERSION TO ELECTRICITY, August 1-3, 1978, Pittsburgh, Pa.

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CHAPTER THREE: ECONOMIC ANALYSIS

This part contains data on the capital and operating costs of the Slagging Lurgi process.

3.1 Introduction and Methodology

3.1.1 Economic Analysis Methodology

The economic analysis relies on a commercial size Slagging Lurgi gasification plant design made by Continental Oil (3-1). The data presented in this report was adjusted for inflation since 1978 when the report was prepared and the adjusted data was used to compute product costs for the process.

3.1.2 Scaling Exponents

The Continental Oil design capacity, 250 billion Btu/day was judged a typical commercial size. Therefore, no scaling was necessary.

3.1.3 Price Indices

Costs presented in the Continental Oil report were presented in first quarter 1978 dollars. Several cost indices were used to correct the cost of capital equipment, other capital costs, and operating costs to third quarter 1980 dollars as was explained in the background section.

3.1.4 Economic Criteria

The standard economic criteria described in the Background section were used. The investment schedule over the four-year construction period was: 10.4 percent, 32.4 percent, 38.6 percent, and 19.5 percent.

3.1.5 Contingencies

Two contingencies were applied to the capital cost estimates: a process contingency and a project contingency. The process contingency covers technical uncertainties within a particular process which might cause costs to increase and was derived judgmentally by ERCO with reference to industry contacts. The percent process contingency applied to each area is shown in Table 3-1. Gasification, methanation, and shift conversion, at the demonstration plant stage, receive a 10 percent contingency. All other areas were judged to be fully commercially developed and received no process contingency.

A project contingency of 15 percent was applied to the total of the costs of each area and unit (not including process contingencies) and contractor's fees. This project contingency is meant to allow for unanticipated cost increases, which usually arise as the plant design is made more complete.

TABLE 3-1

PROCESS CONTINGENCY BY PLANT AREA

NUMBER	ITEM	CONTINGENCY (PERCENT)
100	Coal storage and handling	0
200	Coal preparation	0
300	Gasification and power recovery	10
1200	Raw gas cooling	0
1300	Acid gas removal and gas cleaning	0
1400	Sulfur recovery and tail gas treating	0
1700	Shift conversion	10
1800	Methanation	10
1900	Air separation	0
2000	Utilities and support systems	0
2010	Offsites and miscellaneous	0

3.2 Capital Costs

3.2.1 Itemized Capital Costs

The Total Plant Investment for a 250 billion Btu per day Slagging Lurgi plant would be $1401 million in third quarter 1980 dollars. The largest component of this cost would be Area 2100, Offsites and Miscellaneous, at $304.7 million, or 26.9 percent of the total before fees and contingencies. Area 2000, Utilities and Support Systems, which includes the expensive steam generation and power recovery units, would cost $281.8 million or 24.9 percent of the subtotal. Gasification represents a relatively small portion of the costs, at $58.7 million. Total Plant Investment by area and unit number is presented in Table 3-2.

Besides the capital required to construct the facility, funds are necessary to pay interest on the construction loan and to maintain initial plant operations. Because of the long construction period, interest during construction adds $554.9 million to the capital requirement. Miscellaneous charges include the initial charge of catalysts and chemicals, royalties, starting costs and working capital. They add $239.5 million to the capital requirement. Zero escalation during construction was assumed. The total capital requirement of $2195.4 million is itemized in Table 3-3.

3.2.2 Variability of Capital Cost Estimate

The commercial plant cost estimate made by Continental Oil (3-1) was of sufficient detail to be considered reliable within ±25 percent. The estimate was based on a preliminary equipment list with items sized and specified. Preliminary

TABLE 3-2

TOTAL PLANT INVESTMENT: SLAGGING LURGI[a]

AREA	UNIT	ITEM	COST (10^6 $)[b]	PERCENT OF SUBTOTAL
200		Coal preparation	40.8	3.6
300		Gasification	58.7	5.2
1200		Gas cooling	11.3	1.0
1300		Acid gas removal and gas cleaning		
	1310	H_2S and CO_2 removal	143.9	12.7
	1320	Ammonia recovery	9.7	.9
	1330	Tar and oil separation	15.5	1.4
	1340	Phenol recovery	7.4	.7
1400		Sulfur recovery and tail gas treating	8.4	.7
1600		Product gas compression	16.4	1.4
1700		Shift conversion	64.7	5.7
1800		Methanation	36.0	3.2
1900		Air separation	133.3	11.8
2000		Utilities and support systems		
	2010	Steam generation and power recovery	228.6	20.2
	2020	Wastewater treating and water supply	49.3	4.4
	2030	Solids disposal	3.3	.3
	2040	Plant and instrument air	.6	0
2100		Offsites and miscellaneous		
	2110	Flare and incineration	13.7	1.2
	2120	Tankage, shipping and receiving	6.6	.6
	2130	Other support facilities	284.4	25.1
		Total	1132.6	100
		Contractor's fees	71.7	
		Project contingency	180.7	
		Process contingency	16.0	
		Total Plant Investment	1401	

[a]Source: Reference 3-1, updated by ERCO.
[b]Third quarter 1980 dollars.

TABLE 3-3

SLAGGING LURGI: TOTAL CAPITAL REQUIREMENT[a]

ITEM	UNIT COST (10^6 $)[b]	PERCENT OF TOTAL
Total plant investment	1401	63.8
Escalation during construction	0	0
Interest during construction	554.9	25.3
Miscellaneous		
Initial charge of catalysts and chemicals	20.9	1.0
Royalties/intangible assets/land and land rights	49.0	2.2
Starting costs	84.1	3.8
Working capital	85.5	3.9
Total	2195.4	100

[a]Source: Reference 3-1, updated by ERCO.
[b]Third quarter 1980 dollars.

stream flows, and energy and material balances were worked out. This level of design detail corresponds to midway between an order-of-magnitude and a budget authorization estimate as defined in the Chemical Engineering Handbook (3-2). An order-of-magnitude estimate is made within ± 30 percent, while a budget authorization estimate is within ± 20 percent. Therefore, the cost estimate made in reference 3-1 is accurate within about ± 25 percent.

Although the Slagging Lurgi process has not been tested at commercial scale, technical uncertainties are not a major potential source of cost increases. All sections of the plant except gasification, shift conversion and methanation are composed of commercialized technologies which have been used at the necessary scale. The three untested areas account for roughly 15 percent of the total plant capital requirement. Therefore if the costs of the untested areas were to double, the capital requirement would increase by only 15 percent.

3.3 Operating and Maintenance Costs

3.3.1 Itemized Operating and Maintenance Costs

Annual operating and maintenance (O&M) costs, excluding fuel, are presented in Table 3-4. These costs assume a 90 percent capacity factor. Local taxes and insurance, at $38.42 million are the largest component of O&M costs. Labor amounts to $27.04 million and catalysts and chemicals to $22.15 million. Other charges bring total O&M costs up to $120.3 million.

TABLE 3-4

SLAGGING LURGI:
NET ANNUAL OPERATING AND MAINTENANCE COSTS - 90 PERCENT OPERATING FACTOR

ITEM	ANNUAL COST[b] (10^6 $)	PERCENT OF TOTAL
Gross Annual Operating and Maintenance Costs		
Administration and general overhead	16.90	14.0
Local taxes and insurance	38.42	31.9
Labor		
Process operation	5.63	4.7
Maintenance	17.05	14.2
Supervision	4.37	3.6
Total	27.04	22.5
Supplies		
Operating	1.69	1.4
Maintenance	11.37	9.5
Total	13.06	10.9
Catalysts and Chemicals		
Catalysts	9.77	8.1
Chemicals	12.38	10.3
Total	22.15	18.4
Purchased Water	2.75	2.3
Total Operating and Maintenance	120.3	100
By-Product Credits	(10^6 $)	
Sulfur	(10.8)	
Ammonia	(6.7)	
Tar - Oil	(13.9)	
Phenols	(3.6)	
Naptha	(9.5)	
Total	(44.6)	
Net Annual Operating and Maintenance Cost	(10^6 $)	
Gross Operating and Maintenance Costs	120.3	
By Product Credits	(44.6)	
Total	75.7	

[a]Source: Reference 3-1, updated by ERCO.

[b]Third quarter 1980 dollars.

The Slagging Lurgi process produces by-product sulfur, ammonia, tar-oil, phenols and naphtha. The sulfur and ammonia are sufficiently pure to be sold as chemical feedstocks. The by-product hydrocarbons are not of sufficient purity to allow their sale as substitutes for petroleum based feedstocks (3-3). Instead, they will be sold as fuel. Tar-oil is the most important by-product. Table 3-5 also lists the annual by-product credits, which total $44.6 million. With expenses at $120.3 million, and credits at $44.6 million, net annual operating and maintenance costs total $75.7 million.

3.3.2 Variability of Operating and Maintenance Costs

Variable operating and maintenance costs, including supplies, catalysts, chemicals and purchased water, are essentially a function of the plant design. By-product production, and as a result by-product credits, is also a function of the plant design. Therefore, by-product credits and variable operating and maintenance expenses cannot be considered highly variable as long as the plant design remains static.

Fixed operating and maintenance (O&M) expenses make up the bulk of O&M expenses (78.4 percent) and include labor, administration and general overhead, and local taxes and insurance. These costs were completely itemized in Reference 3-1, and so are probably not subject to large variability due to omissions or overestimations.

In general, the operating and maintenance expense estimate is at least as reliable as the capital cost estimate (± 25 percent).

3.4 Effect of Technology Development on Costs

As the number of Slagging Lurgi gasification plants in service increases, capital costs will decline in real dollars due to the effects of experience. Ten percent has been estimated as the upper limit on the experience factor for new energy process technology (3-4).

The 10 percent experience factor is valid only for the plant costs accounted for by new technology. Most sections of the Slagging Lurgi plant employ mature technologies whose costs would decline little as more Slagging Lurgi plants were built. The accumulated volume of production of these mature technologies is so large that the construction of one or several Slagging Lurgi plants would result in small additional cost reductions because of experience. Novel components, including gasification and power recovery, raw gas cooling, shift conversion and methanation, account for about 15 percent of the total plant investment. Therefore, the experience factor for Slagging Lurgi technology would be 15 percent times 10 percent, or about 2 percent. Each doubling of Slagging Lurgi production capacity would result in a 2 percent reduction in unit capital costs.

3.5 Gas Costs

The cost of the product gas has three components: capital charges associated with plant capital costs, plant operating and maintenance (O&M) costs, and coal costs. The cost of the gas excluding the cost of coal (non-fuel costs) indicates the cost of converting the coal to clean gas. Non-fuel gas costs can be computed from capital charges and O&M costs according to the formula given in the Background.

This formula yields a non-fuel gas price of:

$$P = \frac{(\$2195.4 \times 10^6 \times 20\ \%) + \$75.7 \times 10^6}{(91.25 \times 10^{12} \times 90\%)}$$

$$= \underset{\text{(Capital costs)}}{\$5.35/10^6\ \text{Btu}} + \underset{\text{(O\&M costs)}}{\$.92/10^6\ \text{Btu}} = \underset{\text{(Total)}}{\$6.27/10^6\ \text{Btu}}$$

Capital costs are $\$5.35/10^6$ Btu, and operating and maintenance costs are $\$.92/10^6$ Btu, for a total non-fuel gas cost of $\$6.27/10^6$ Btu.

The non-fuel gas cost can be converted to a total energy cost by adding the cost of coal as was explained in the Background. The Slagging Lurgi process has an overall thermal efficiency of 59.1%. Therefore, the fuel component of the gas would cost $\$2.54/10^6$ Btu with coal at $\$1.50/10^6$ Btu. The total energy cost is $\$8.81/10^6$ Btu. Note that the capital and O&M cost estimates are accurate within only $\pm 25\%$.

References

3-1. Continental Oil Company, "Phase I: The Pipeline Gas Demonstration Plant, Volume 3. Part One: Economic Analysis. Part Two: Technical Assessment." United States Department of Energy, FE-2542-10 (Vol. 3), 1978.

3-2. Perry, Robert H. and Cecil H. Chilton, eds. Chemical Engineering Handbook, Fifth Edition (New York: McGraw-Hill, 1973), pp. 25-12 - 25-17.

3-3. United Technologies Research Center, "Gasification Combined-Cycle Power Generation: Environmental Assessment of Alternative Systems." Argonne National Laboratory, 1978, Number ANL/ECT-7, pp. 195-202.

3-4. Hederman, W.F. (Rand Corporation). "Prospects for the Commercialization of High-Btu Coal Gasification." U.S. Department of Energy, April, 1978, Number R-2294-DOE, pp. 48-50.