5.2.2 BGC/Lurgi Utilization in a SASOL Type Plant

5.2.2.1 Synthesis Gas Production

The substitution of the BGC/Lurgi gasifier would not alter the basic process flow of the SASOL type plant shown in Figure 4-1

Figure 5-3 shows the process flow in the gas preparation section. Fourteen gasifiers (including two standby units) would be needed to replace the 28 Lurgi gasifiers used in the SASOL-U.S. system designed by Mobil. For a comparable total coal input, throughput to the BGC would be 2070 M lbs/hr as opposed to 1900 M lbs/hr for the Dry Ash Lurgi as a result of the reduction in coal diverted to meet steam requirements. The gasifier steam requirement would be reduced from 1700 M lbs/hr to 543 M lbs/hr. An additional 252 M lbs/hr is required for the shift reaction. The material balances for the resulting gasification and shift reaction are shown in Table V-2. After shift and purification the H_2/CO of the gas produced by the BGC design is similar to the base case (Reference Table 4-2). As a result of the lower methane content and higher efficiency (e.g., lower steam consumption) of the BGC design, the moles of H_2 and CO are 45% greater than in the Base Case.

5.2.2.2 Plant Products and Efficiency

Table V-3 compares products produced after F-T synthesis and upgrading in the base case with those from BGC gasification for both mixed product and all liquid cases. In the mixed product case BGC gasification results in about 17% less SNG production, and about 45% more

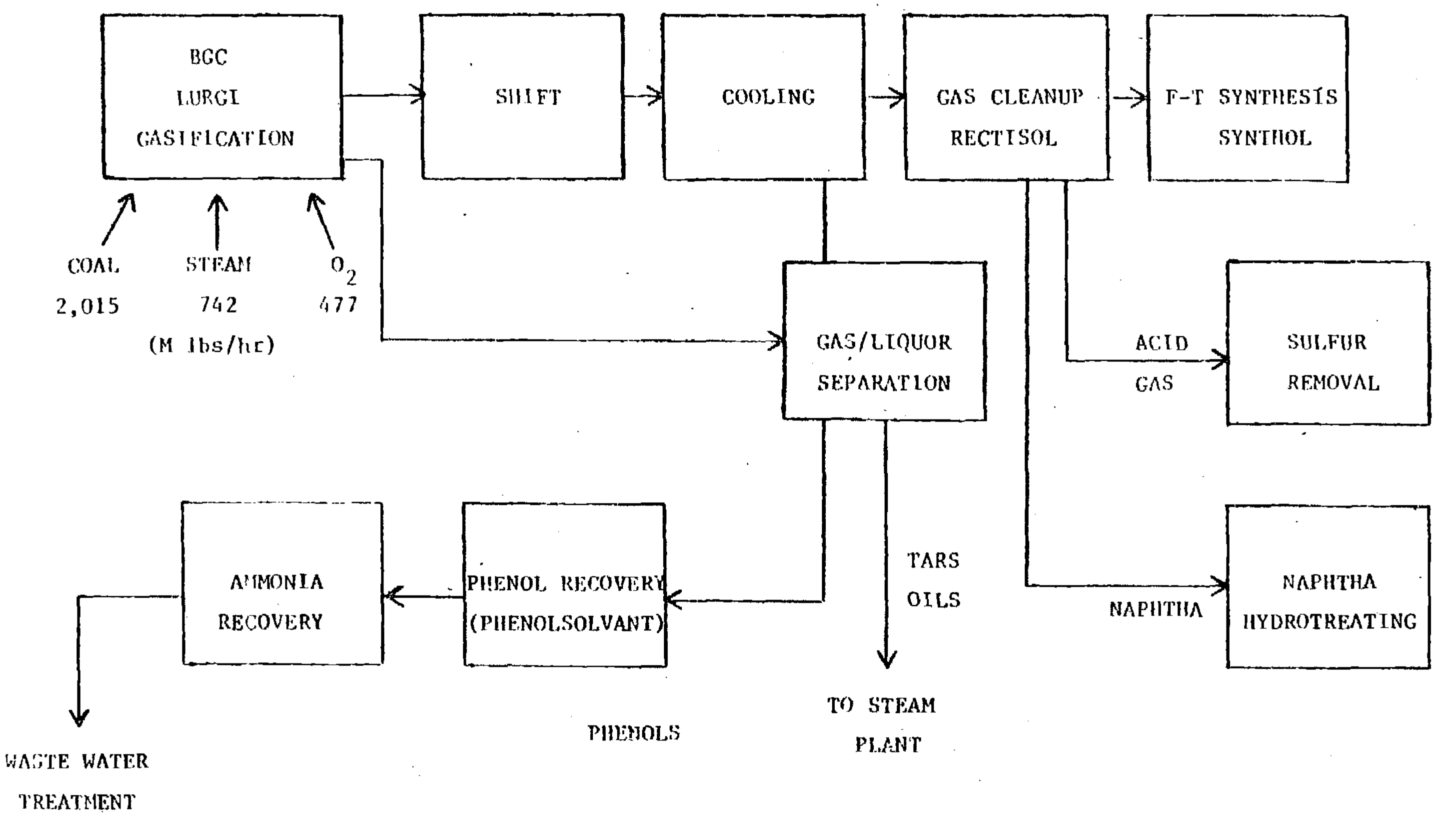

FIGURE 5-3
SYNTHESIS GAS PREPARATION BGC/F-T

TABLE V-2

MATERIAL AND ENERGY FLUX IN SYNTHESIS GAS PREPARATION UNITS,
BGC GASIFICATION, SYNTHOL SYNTHESIS

	BGC Gasifier				Cool Shifted Gas		Purified Gas	
	Input 1000/hr		Output 1000/hr		Output 1000/hr		Output 1000/hr	
	lb-mole	lb	lb-mole	lb	lb-mole	lb	lb-mole	lb
CO_2			3.0	132	39.8	1751	4.5	197
CO			63.8	1786	27.0	757	26.8	750
H_2			31.9	64	68.7	138	68.1	137
CH_4			7.5	121	7.5	121	7.5	120
C_2H_4/C_2H_6			.7	23	.7	23	.4	10
H_2O			38.3	689	.2	6	nil	nil
Misc. Gases			.5	13	.5	13	.3	8
Naphtha				17		17		
Tars/Oils/Phenols				120*				
Coal/(Ash)		2070		105				
Steam		453						
Oxygen		545						
Totals		3068		3068	144.4	2826	107.6	1222
Energy (LHV) MM Btu/hr	16944 (DAF COAL)		16278		13830		13116	

*Includes Tars and Phenols recovered from quench water

TABLE V-3

COMPARISON OF PRODUCTS AFTER DOWNSTREAM PROCESSING
(BGC/SYNTHOL AND SASOL-U.S. FOR MIXED AND ALL-LIQUID OUTPUT)

	Mixed Output		All-Liquid	Output
	SASOL-U.S.	BGC	SASOL-U.S.	BGC/SYNTHOL
SNG MM SCF/D	173.3	147.9	-	-
Gasoline B/D	13,580	19,137	28,090	31,514
C_3 B/D	1,107	1,604	2,436	2,738
C_4 B/D	146	212	321	361
Deisel B/D	2,307	3,343	5,078	5,706
Fuel Oil B/D	622	901	1,369	1,538
Alcohol B/D	1,829	2,650	4,026	4,524
Total Liquids B/D	19,591	27,847	41,230	46,381
FOE B/D	44,950	47,418	33,652	37,776
Efficiency (HHV)	57	60.1	42.7	47.9
Liquid Fuels/Ton Dry Coal (B)	.98	1.37	2.07	2.32
Liquid Fuels C_4^+/Ton Dry Coal (B)	.92	1.31	1.94	2.18

production of liquid products. Overall efficiency is increased from 57.3% to 60.1%. The reason why the overall efficiency improvement does not reflect the full 10% improvement in net gasifier efficiency is illustrated in Figure 5-4, which shows the system energy losses for the plant employing the BGC/Lurgi. Note the overall efficiency of the synthesis section is 82.6 versus 86.8 for the Base Case SASOL-U.S. plant (Reference Figure 4-7). This decreased efficiency occurs because a higher percentage of the total output is converted in the Synthol unit where the efficiency is 79%, and a lower percent follows the SNG train where conversion losses are minimal. However, even with the more favorable liquid output, over half of the total plant output is SNG. Gasifier modification alone cannot produce a yield which is predominantly liquid, since 25% of the Synthol reactor output is C_1 and C_2, and the unconverted synthesis gases (15% of input) are methanated. SNG in a mixed output plant with a design similar to the SASOL-U.S. Base Case would thus approach 40% of total output if <u>no</u> methane were produced in the gasifier.

The advantages of the BGC gasifier are somewhat magnified if used in a plant designed for an all liquid output.. These outputs, also shown in Table V-3 result in product yield more than 12% greater than was achieved with Dry Bottom Lurgi gasification. This improvement stems from the better gasifier efficiency, and from the lower methane make which results in less losses when the methane is reformed to

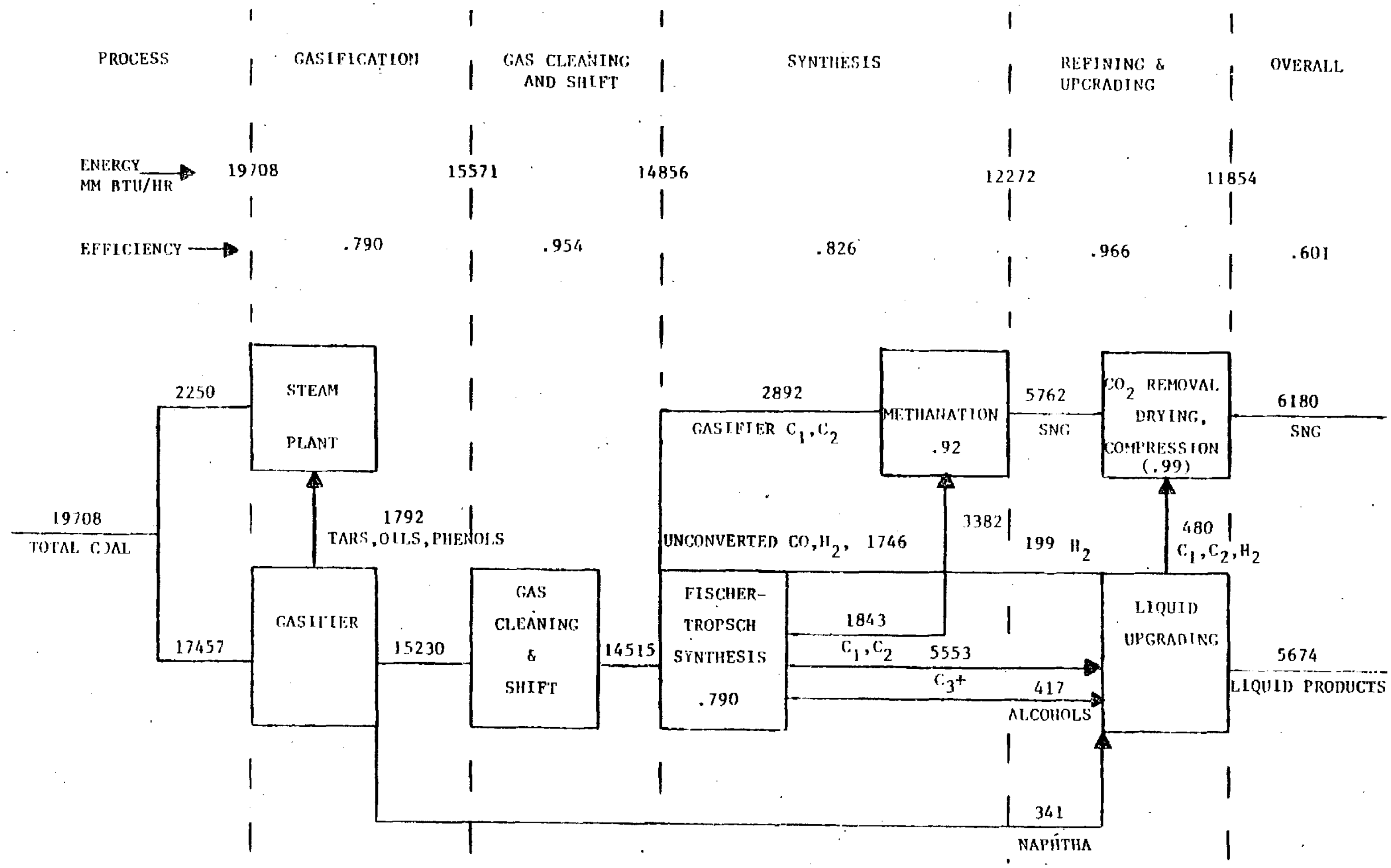

FIGURE 5-4
SYSTEM ENERGETICS FOR BGC GASIFICATION/ SYNTHOL SYNTHESIS MIXED PRODUCT CASE

produce additional synthesis gas. The net result is an overall efficiency of coal to liquid products of 47.9%. With this efficiency, the barrels per ton of dry coal exceeds 2.3 and it thus approaches the values obtained with direct hydrogenation processes. Furthermore, the output products in this instance are fully refined products meeting current consumer specifications, whereas indirect liquefaction products usually require upgrading to be useful as other than boiler fuel.

5.2.2.3 Cost Impact of BGC Gasification

Table V-4 compares plant construction costs for the base case and for a similar plant with BGC gasification. Major savings in the plant employing BGC Lurgi gasification are in gasifier cost and a lower cost steam plant. These savings are partially offset by increased shift costs, and the requirement for larger F-T synthesis and upgrading facilities. The total construction cost saving is seen to be $71.8 M or about six percent in the mixed product case, and $93.3 M or seven percent in the all liquid case. Details of the construction cost analysis are presented in Appendix C.

Table V-5 shows a comparison of the cost of products produced by the base case plant and the alternative plant using BGC Lurgi gasification. These projections are based on the assumption that all cost elements other than coal costs are proportional to plant construction cost. Gasoline costs are computed assuming that all products are sold at the same value per MMBtu (e.g., Thermal Basis) and assuming that products other than gasoline have the defined market

TABLE V-4

CONSTRUCTION COST COMPARISONS
Base Case SASOL-U.S. & BGC/Synthol
$M (1977)

	Mixed Output		All Liquid Output	
	Base Case	BGC Synthol	Base Case	BGC Synthol
Coal & Ash Handling	71.4	71.4	71.4	71.4
Steam Plant	195.3	156.4	212.9	172.8
Oxygen Plant	110.1	124.3	148.5	155.5
Gasification	200.7	100.4	200.7	100.4
Shift	12.8	30.0	12.8	30.0
Gas Cooling & Cleaning	118.1	127.9	118.1	127.9
Sulfur Recovery	59	64.8	59	64.8
Gas/Liquor Separation & Product Recovery	51.4	15.5	51.4	15.5
Waste Water Treatment	26.3	16.3	26.3	18.4
F-T Synthesis	76.4	99.0	109.1	147.7
F-T Product Upgrading*	128.9	154.4	172.1	186.9
F-T Catalyst Preparation	27.7	35.9	48.0	53.6
Auto Thermal Reformer			40.7	36.5
Miscellaneous	108	108	108	108
TOTAL	1186.1	1104.3	1382.7	1289.4

*Includes methanation and SNG preparation, where applicable.

TABLE V-5

GASOLINE COSTS FOR MIXED PRODUCT
AND ALL LIQUID SASOL-U.S. PLANTS

	Mixed Output		All-Liquid Output	
	SASOL-U.S.	BGC/Synthol	SASOL-U.S.	BGC/Synthol
THERMAL PRODUCT BASIS				
$/MMBtu	7.78	6.91	11.95	9.99
$/Gal	.93	.83	1.43	1.19
MARKET BASIS*				
$/Gal	1.33	.92	1.51	1.24

*Assumes products other than gasoline sold at prices given in Table IV-7 values given in Table IV-7 (e.g., Market Basis). Computational details are presented in Appendix D.

In the mixed product case, the combination of reduced plant cost and higher yield result in gasoline costs per gallon of $.83 on a Thermal Basis and $0.92 on a Market Basis, as compared to values of $.93 and $1.33 respectively for the SASOL U.S. Base Case plant. Gasoline cost reductions made possible by the BGC gasifier are somewhat greater in the all liquid case where Thermal Basis and Market Basis costs are $1.19 and $1.24 respectively compared to $1.43 and $1.51 for the Base Case plant using Dry Bottom Lurgi gasification. As in the base case, the modification required to produce an all liquid product are not cost effective if SNG can be sold for $6.17/MM Btu.

5.2.2.4 Environmental Considerations

The primary advantage of the BGC from an environmental point of view results from the character of the ash. Ash from the BGC is a fused, non-leachable "frit" which should pose minimal disposal problems compared to the semi-soluble powder recovered from the Dry Ash Lurgi.

Other improvements offered by the BGC stem from the lower levels of pollutants generated. Waste water requiring treatment is reduced by a factor of 2. The higher overall efficiency of plants employing BGC gasification would be reflected in corresponding lower levels of CO_2 generation. Similarly, higher efficiencies reduce the amount of mining and consequent environmental damage which must be incurred per unit of output product.

5.2.2.5 Conclusions

The improvements in construction cost and output made possible by use of BGC gasifiers are promising; particularly in the case where all liquid outputs are desired. Furthermore, the advantages shown are believed to be a conservative statement of the advantages of the BGC gasifier vis-a-vis the Dry Bottom Lurgi for this application. Steam and oxygen requirements for BGC gasification of Western coal have been conservatively estimated. Further conservatism results because minimal changes were made in the design and operation of the overall plant in order to permit a simple and direct comparison to the Base Case. Plant modification tailored to BGC capabilities could only improve the relative results.

5.3 Texaco and Shell-Koppers Gasifiers

5.3.1 Entrained Flow Gasifiers

The entrained flow gasifier operates with the pulverized or ground coal and the gasifying medium (oxygen and steam) fed co-currently. Figure 5-5 shows a schematic of entrained flow gasification. The advantages are:

- Processes entire mine output (i.e., can handle coal fines)
- Processes caking and non-caking coal; no pre-treatment required to process caking coals.
- Coal residence time is short and gasifier throughput is relatively high
- High carbon utilization

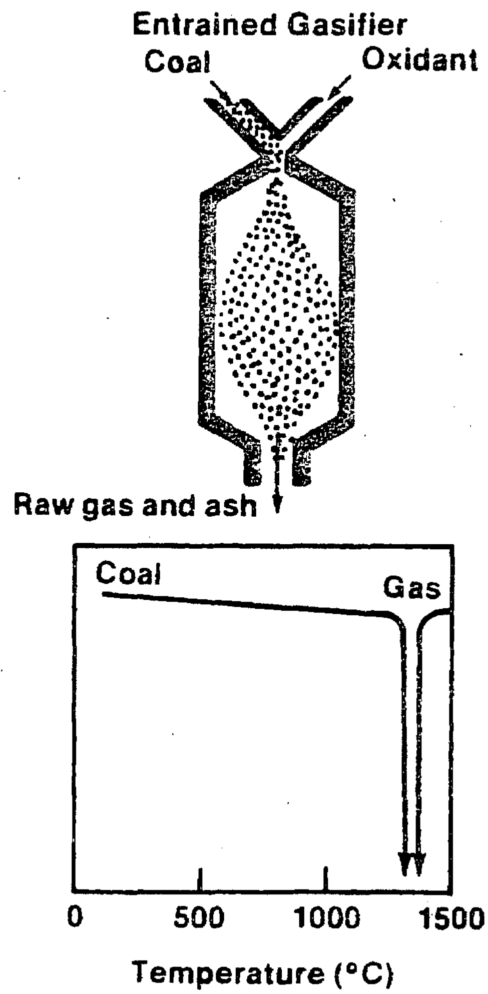

FIGURE 5-5
ENTRAINED SUSPENSION COAL GASIFICATION

- No tars and minimal methane

- Excellent environmental compliance as regards emissions and solid wastes

Disadvantages are:

- Higher moisture coal requires drying for maximum gasification efficiency

- The low H_2/CO molar ratio raw gas product requires external CO water shift to achieve the higher ratios required for purified synthesis gas

- Recovery of the sensible heat from the gasifier product is required if a high net thermal efficiency is to be achieved. The design of waste heat boilers to recover this heat is difficult because of the high temperature and molten as present.

5.3.2 Operations Experience of Texaco and Shell-Koppers

The Texaco and Shell-Koppers entrained flow gasifiers have had satisfactory experience in pilot and demonstration plants. Texaco's coal gasifier pilot plant research is carried out at Montebello, California and the present development effort commenced soon after the global oil crisis of 1973.[16] The pilot plant's two gasifiers are capable of processing 15 to 20 tons per day of coal. Tests have been run on a wide range of coal at pressures ranging from 300 to 1200 psi. In 1978, a 150 TPD coal gasification demonstration plant was started at the Ruhrchemie Chemical Plant complex in Oberhausen-Holten, West Germany as a joint venture by Ruhrchemie A.G., Ruhrkohle A.G. and the West German government. In 1979, the TVA started to convert a natural gas fueled ammonia plant at Muscle Shoals to coal by installing a 150 TPD Texaco coal gasifer.[17]

Start up was planned for August 1980. Texaco and Southern California Edison Company in 1979 announced a joint venture involving Electric Power Research Institute (EPRI) funding to construct a 1000 TPD coal gasification demonstration plant to supply medium-Btu fuel gas to a 1000 MW combined cycle demonstration plant at SCEC Cool Water plant near Daggett, California. Operation is scheduled for late 1983. The Tennessee Eastman Company announced in 1980 that they will install a Texaco gasifier (estimated 1000 TPD) to produce synthesis gas for the manufacture of methanol and acetic anhydride. Operation is scheduled for mid-1983.

Shell-Koppers started in December 1976 with a 6 TPD gasifier pilot plant, located in Amsterdam, Holland. In November 1978, they started a 150 TPD coal gasifier at Deutsche Shell's Harburg, West Germany petroleum refinery.[19] Design work is now underway (1980) for two 1100 TPD prototype coal gasification plants. One plant will be built at Moerdijk, Netherlands; the second will be built in West Germany. The two plants are scheduled to be in operation in 1984 and 1985, respectively. Shell-Koppers is planning on progressively larger capacities up to a maximum of about 2750 TPD coal by the late 1980's.[20]

5.3.3 Interpreting Texaco and Shell-Koppers Published Data

Both Texaco and Shell-Koppers gasifier developments are self-financed and the information developed is considered highly confidential and of great commercial value. Therefore, very little is known. Some Texaco information has been published.[17,19] Other information can be derived from various EPRI and DOE sponsored reports such as the excellent Fluor Corporation report for EPRI in 1978.[7] Shell-Koppers data are limited to two sources;[20,21] the best source being that published in CEP, March 1980.

Use of the data is conditioned on the following assumptions:

- Extrapolation of pilot plant data is based on assuming

  (1) the published gasifier thermal efficiency, and
  (2) the published H_2/CO molar ratio in the raw gas effluent

- Pilot plant data are extrapolated "stoichiometrically" to determine

  - Oxygen and steam requirements
  - CO_2 and water content in raw gas
  - Hydrogen balance

However, once the thermal efficiency and H_2/CO molar ratio are assumed, the extrapolation is severely constrained. Stoichiometric relationships are considered to be only guidelines. The Texaco Development Corporation, Harrison, New York, the licensors of the Texaco Process, repeatedly stated in response to inquiries for information details that, "it depends on the coal characteristics;" and, with that, refused to supply any more information than was available in the literature. The Shell-Koppers developments are relatively recent, and inquiries for additional information have been rejected. Therefore, resort was made to qualitative and quantitative information on general gasifier studies, such as the excellent Shinnar analysis.[13]

The reactions and reaction rates within a gasifier are not precisely known. There have been many studies published on the stoichiometry, reaction kinetics, and thermodynamics involved, as the excellent analyses by Shinnar,[13] and Wei.[22] But, understanding gasifier operations and extrapolating pilot plant data must still be done on an empirical basis derived from pilot plant work. Theoretical developments do not currently lead to design. Most importantly, the type of coal and its characteristics make it almost certain that any Texaco or Shell-Koppers gasifier process design, must be preceded by careful and detailed pilot plant work.

In summary, analysis of Texaco and Shell-Koppers gasifiers depends on data published for a specific coal. Application of the analysis to other coals should be considered as less rigorous in accuracy and subject to confirmation by pilot plant tests performed on the specific coal under consideration.

5.3.4 <u>Critical Evaluation Factors for Texaco and Shell-Koppers</u>

Currently, there is no commercial operations experience with Texaco and Shell-Koppers gasifiers. Therefore, there is now no definitive procedure to deal with the following uncertainty areas:

- Coal drying for Texaco and Shell-Koppers
- Pressurized dried coal feeder for Shell-Koppers
- Coal slurry for Texaco

This report is based on processing 27,800 TPD of a Wyoming subbituminous coal whose specifications are outlined in the MRDC

Report. A key feature of this coal is the very high inherent moisture content, namely, 18%. The Texaco gasifier relies on a coal slurry feed as the means by which the coal feed is accurately metered, controlled, and injected. According to Texaco, the maximum coal solids content that may be carried in a water slurry is 70% by weight. Since the Wyoming coal already contains 28% moisture, the amount of carbon available is severely reduced if it is necessary to slurry as received coal. If the 28% moisture coal is slurried 70/30 coal : water, the resulting moisture to dry coal ratio is about 50/50. If the coal is first dried to 5% moisture the ratio of coal : water is improved to 66/34. However, there is the additional problem of moisture re-adsorption by the dried coal in the slurry. Depending on the physical characteristics, including importantly, the particle size, the amount of water reabsorption from the slurry water may increase the effective solid/liquid ratio to a point where additional slurry water is required. This reduces the percentage coal and severely reduces the efficiency of the Texaco gasifier by increasing the amount of water that has to be vaporized.

The Shell-Koppers unit gasifies coal by partial oxidation. The gasifying medium is oxygen. A relatively small amount of steam is added merely to achieve the proper hydrogen stoichiometric balance. Therefore, it is essential that the coal be dried to a low level. In the Shell-Koppers literature,[20] the coal is dried to as low as 1% moisture in the Harburg, West Germany 150 TPD gasifier. As indicated in Section 5.3.6, drying the Wyoming sub-bituminous coal to as low as 5% inherent moisture presents technical problems of varying severity and complexity.

The key mechanical feature of the Shell-Koppers gasifier, as compared to Texaco, is the presence of the pressurized (nitrogen gas) dried coal feeder. This is an important technical advance because the coal slurry feed principle is eliminated. The reduction in total water (coal inherent moisture plus slurry water) which is injected into the gasifier is essential to achieving the very high thermal efficiency (83%) and carbon utilization (99%) of the Shell-Koppers gasifier as compared to the efficiency of the Texaco (77%). There exists some industry viewpoints that a mechanically satisfactory long-life dried coal feeder is not yet commercially available. If this viewpoint were confirmed, then the Shell-Koppers gasifier operation, as now designed, would not be possible. According to Shell-Koppers, the pressurized dried coal feeder works satisfactorily at Harburg and there should be no problem with commercial design and fabrication of the feeder.[20]

The Texaco and Shell-Koppers gasification and subsequent waste heat recovery and raw gas purification are similar. Therefore, evaluation of the processes will be made on a similar basis.

The other major category of uncertainty is the steam balance, According to the literature, the waste heat recovery available from both Texaco and Shell-Koppers is sufficient to produce steam to drive the air compressor and oxygen gas compressor steam turbine drives in the oxygen plant. This is a key advantage factor attributed to entrained gasifiers. And, it is confirmed by pilot plant data. However, in the indirect liquefaction plant complex, it is necessary to have an autothermal SNG reformer, employing oxygen gas to reform the methane, ethylene, and ethane arising from the Fischer-Tropsch reactor and the liquid products upgrading. This reforming step is a partial oxidation. The oxygen required increases oxygen plant capacity by about 15%, and therefore, turbine drive steam. The steam balance then becomes of utmost criticality to achieve a satisfactory overall plant thermal efficiency. Therefore, waste heat recovery and energy conservation practice in the gasification plant becomes of paramount importance.

5.3.5 Indirect Liquefaction with Texaco and Shell-Koppers Gasification

Figure 5-6 is a flow diagram of the synthesis gas preparation section of an indirect liquefaction plant employing Texaco gasifiers. The major difference from the SASOL-U.S. base case is the elimination of a separate steam generation unit. Steam required for powering the air separation plant is produced from heat recovered from the gasifier exit stream in waste heat boilers.

Process flow with Shell Koppers gasification would be similar to Texaco except that the pulverized coal is not required to be slurried, and a nominal amount of steam from the waste heat boiler is introduced in the gasifier.

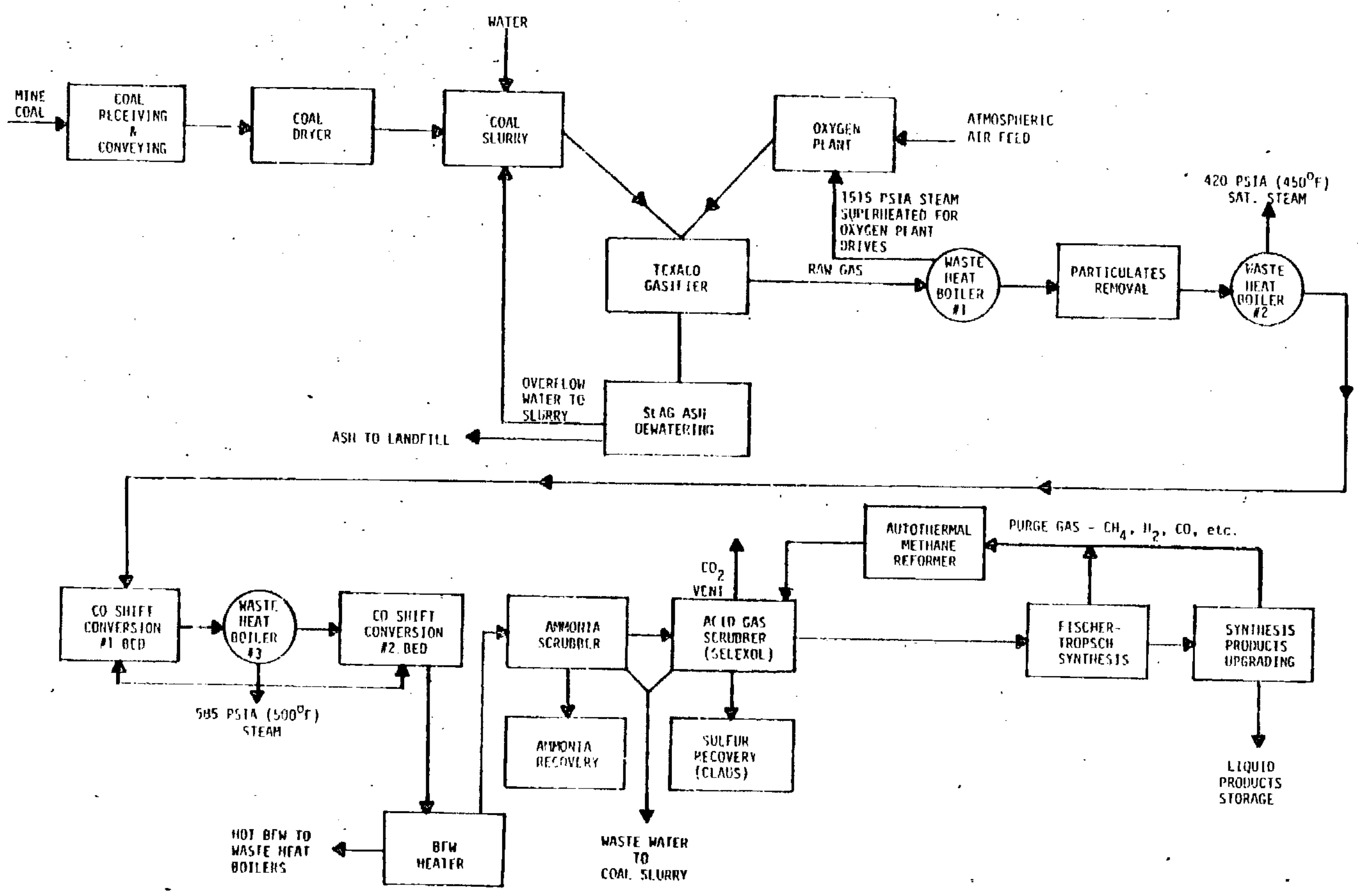

FIGURE 5-6
INDIRECT COAL LIQUEFACTION TEXACO GASIFICATION & FISCHER-TROPSCH PROCESS DIAGRAM

a. Gasifer Vessel

The Texaco gasifier is illustrated in Figure 5-7 as a sketch. The Texaco Development Corporation has refused disclosure of details on proprietary grounds. Details on the auxiliary equipment, namely: coal grinding and water slurrying and the waste heat boiler system (i.e., raw gas cooling) are not available for the same reason. A literature search has produced very little information.

Very much less information is available on the Shell-Koppers unit, perhaps because development has been later and the largest unit is the 150 TPD demonstration plant at Harburg.

Texaco has a design for a 2000 TPD gasifier, 9 feet diameter x 15 feet high.[23] The coal slurry and oxygen gas will be fed co-currently and downward into the cylindrical unit. Current largest gasifiers will be the 1000 TPD unit for the joint venture, Southern California Edison Company-Texaco-EPRI, at Daggett, California, scheduled for late 1983; and, the estimated 1000 TPD unit for Tennessee Eastman Company scheduled for mid-1983 at Kingsport, Tennessee.

The Shell-Koppers unit is an adaptation of the Koppers-Totzek gasifier. The coal, oxygen, and steam will be fed co-currently through two conical feed heads located in the lower portion of the cylindrical gasifier. Design work is now underway (1980) for two-1100 TPD prototype gasifiers; one at Moerdijk, Netherlands for 1984 and the second in West Germany sometime in 1985. The maximum design capacity will be about 2750 TPD scheduled for the late 1980's.

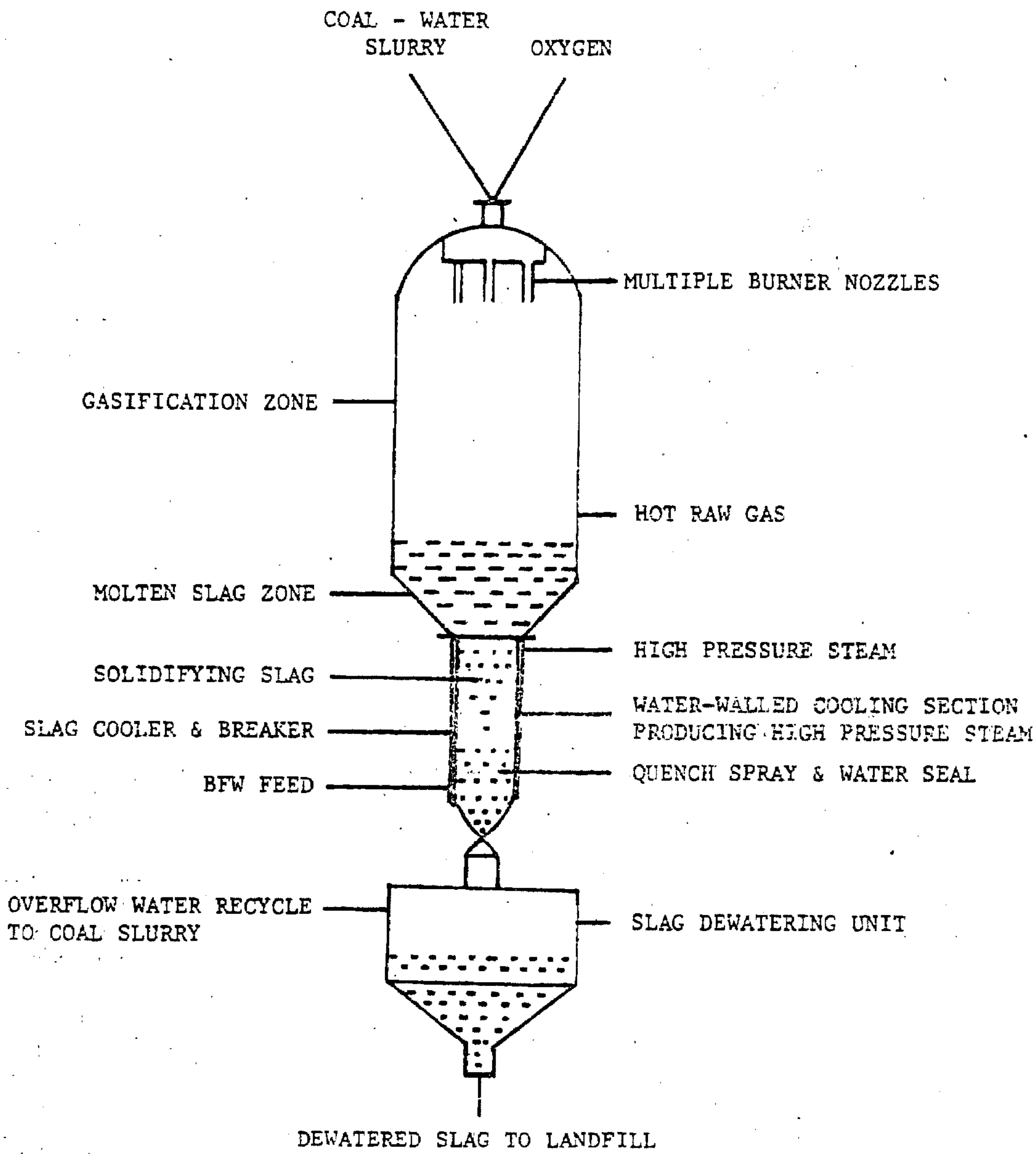

FIGURE 5-7
TEXACO GASIFIER SCHEMATIC DIAGRAM

b. Coal Receiving and Drying

Twenty-seven thousand eight hundred TPD mine coal is received and conveyed by current technology coal-handling equipment. The mine coal is pulverized to size ½" x 0 for the fluidized coal dryer.

There will be fourteen (14) 150 TPH fluidized coal dryers, which will dry the Wyoming sub-bituminous coal from 28% wt. inherent moisture to 5.0% wt. The coal dryers operate on a drying gas stream produced by combustion of 2100 TPD coal. The resulting hot flue gas is tempered flue gas at 900°F raises the temperature of the coal to 225°F while removing the inherent moisture to a dried coal content of 5% wt. The dried coal is conveyed to silo storage and removed as required.

This sequence of events is similar for both Texaco and Shell-Koppers.

c. Dried Coal Grinding, Slurrying and Feeding

In the Texaco process, the dried coal is wet ground in the slurry water and the coal slurry consisting of 30% water and 70% dried coal still containing 5% moisture is prepared. The slurry is temporarily stored or fed directly to the gasifier.

Texaco keeps confidential the details on particle size, grinding and slurrying. They consider the details to be a part of the process license package. Section 8.3 gives details of the sensitivity of Texaco gasifier performance to slurry concentrations lower than the 70/30 ratio assumed here.

Shell-Koppers transports the dried coal pneumatically with nitrogen gas to the coal grinder or mill. The coal is reduced in particle size to "dust."[20] This "dust" is pneumatically fed to the pressurized dried coal feeder. There is a coal flow meter on the pneumatic line which is safety-interlocked to the oxygen gas feed. In the event of variation or disruption of coal feed, the oxygen gas supply is controlled or cut-off, as required.

In the Texaco unit, the coal slurry is concurrently fed with the oxygen and steam. Flow control problems are simplified because the three streams are fluids. The arrangement of feed nozzles in the gasifier top head are proprietary.

d. Gasifier Operation

The Shell-Koppers and Texaco gasifier operations are summarized in Table V-6.

Table V-6 indicates the similarities of operation of entrained gasifiers, but also points out sharply how differently the Shell-Koppers unit compares to the Texaco, namely:

- Shell-Koppers is almost completely a partial oxidation reaction. There is very little steam available for the water gasification reaction. As a derivative of this situation, the Texaco H_2/CO molar ratio is nearly 42% greater than Shell-Koppers because the water gasification reaction is a significant factor in hydrogen production.
- The oxygen demand by Texaco is 24% greater than Shell-Koppers. The additional oxygen is needed to burn coal to vaporize the excess slurry water to steam. Hence, a penalty is paid by Texaco in the form of a larger oxygen plant and a lower gasifier thermal efficiency.

Table V-7 is a self-explanatory in indicating the influence of slurry water excess in promoting a higher H_2/CO ratio for Texaco, whereas in Shell-Koppers, the H_2/CO ratio is much lower because the

TABLE V-6

GASIFIER OPERATIONS

	UNIT	TEXACO	SHELL-KOPPERS
GASIFIERS			
Total Number		14	14
Capacity, Each	TPH	2,000	2,000
THERMAL DATA, LHV, MAF COAL			
Coal Input - Total	MM Btu/Hr	18,959	18,959
Coal Input - Each Gasifier	MM Btu/Hr	1,354	1,354
Raw Gas Output - Total	MM Btu/Hr	15,178	15,735
Net Thermal Efficiency	%	77.0	83.0
STOICHIOMETRY			
H_2/CO Molar Ratio	Mols/Mol	0.68	0.48
99.5% Oxygen/MAF Coal	Lbs/Lb	1.025	0.828
*Steam/MAF Coal	Lbs/Lb	0.486	0.030
OPERATION CONDITIONS			
Steam Supply - Pressure	PSIA	Slurry Water	475
- Temperature	^oF	Slurry Water	500
Oxygen Supply - Pressure	PSIA	705	475
- Temperature	^oF	300	400
Raw Gas Exit - Pressure	PSIA	615	435
- Temperature	^oF	2,600	2,732

* Steam addition; does not include 5.0% Wt. Inherent Moisture in the dried coal.

TABLE V-7

GASIFIER RAW GAS EFFLUENT

	TEXACO		SHELL-KOPPERS	
Component	Mol %	Lb. Mols/Hr.	Mol %	Lb. Mols/Hr.
H_2	27.62	50,000	30.91	44,020
CO	40.62	73,675	64.39	91,700
CO_2	12.23	22,185	2.40	3,415
CH_4	0.08	150	None	None
H_2S	0.12	210	0.15	210
COS	50 PPM	10	50 PPM	10
NH_3	0.45	820	0.57	820
N_2	0.07	120	0.03	120
Ar	0.05	80	0.06	80
H_2O	18.87	34,030	1.43	2,035
TOTAL	100.00	181,380	100.00	142,410
H_2/CO Molar Ratio:	0.68		0.48	