4.0 SASOL-U.S. (BASE CASE)

4.1 Introduction

Much of the detailed technical and economic data on SASOL is proprietary. MITRE has therefore employed a detailed technical and economic description of a SASOL-type plant developed by Mobil Research and Development Corporation with assistance from Lurgi as a Base Case of a SASOL-type indirect liquefaction plant representative of current commercial technology. The MRDC work was performed under DOE contract and is reported in detail in Reference 1. Plant description and other data from the Mobil design are hereafter referred to as SASOL-U.S. or as the Base Case.

4.2 System Description

The SASOL-U.S. design is slightly larger than SASOL II, having a fuel oil equivalent output of about 45,000 barrels/day. It differs further in that it provides for co-production of synthetic natural gas (SNG). Current evidence is that SASOL II will convert all methane produced to synthesis gas to be used for the production of additional liquid hydrocarbons.

The SASOL-U.S. design is based upon the use of a Western U.S. subbituminous coal. Properties of this coal, and the coal to be used in SASOL II are shown in Table IV-1. These subbituminous coals differ in that the SASOL coal is high in ash and relatively low in moisture, whereas the Western U.S. coal is low in ash but very high in moisture. Both are non-caking coals well suited for processing in Lurgi gasifiers.

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TABLE IV-1

COAL CHARACTERISTICS

	SASOL II[6]	WESTERN U.S. (Wyoming Sub-Bituminous)
Proximate:	(Percent by Weight)	
Moisture	5.5	28.0
Ash	25.5	5.1
Volatile matter	23.6	33.1
Fixed carbon	45.4	33.8
Ultimate DAF:		
Carbon	78.7	74.45
Hydrogen	4.2	5.10
Sulfur	1.6	.45
Nitrogen	1.9	.75
Oxygen	13.6	19.25

Figure 4-1 shows a simplified SASOL-U.S. process flow diagram. The five principal steps in the process flow are coal preparation, gasification, clean gas preparation, synthesis, and synthesis product upgrading. The simplified diagram shows two alternative methods of handling purge gas from the Synthol Fischer-Tropsch synthesis unit. The purge gas is a mixture of methane, C_2 hydrocarbons, unconverted hydrogen and CO. This gas may be sent to a methanation unit, wherein the H_2 and CO are converted to methane to produce an SNG product. Alternatively, the purge gas may be processed in a reformer in which methane and C_2 hydrocarbons are reformed to yield additional H_2 and CO, which is then recycled to the Synthol reactor. As noted above, the Mobil design assumes that an SNG co-product is produced, whereas SASOL II will reform the methane to yield additional synthesis gas. Both options are considered in this evaluation.

4.2.1 Coal Preparation

Coal preparation consists of grinding and screening. Particles well sized for Lurgi gasifiers, (e.g., .25 x 2.0 inches), are so used. It is assumed that 82 percent of the coal is suitable for Lurgi gasification. The remaining 18 percent which are too fine for use in Lurgi gasifiers are fired for steam generation.

4.2.2 Preparation of Synthesis Gas

Synthesis gas preparation is shown schematically in Figure 4-2. Material and energy fluxes at the input and output of the gasifier, after shift and after purification are shown in Table IV-2.
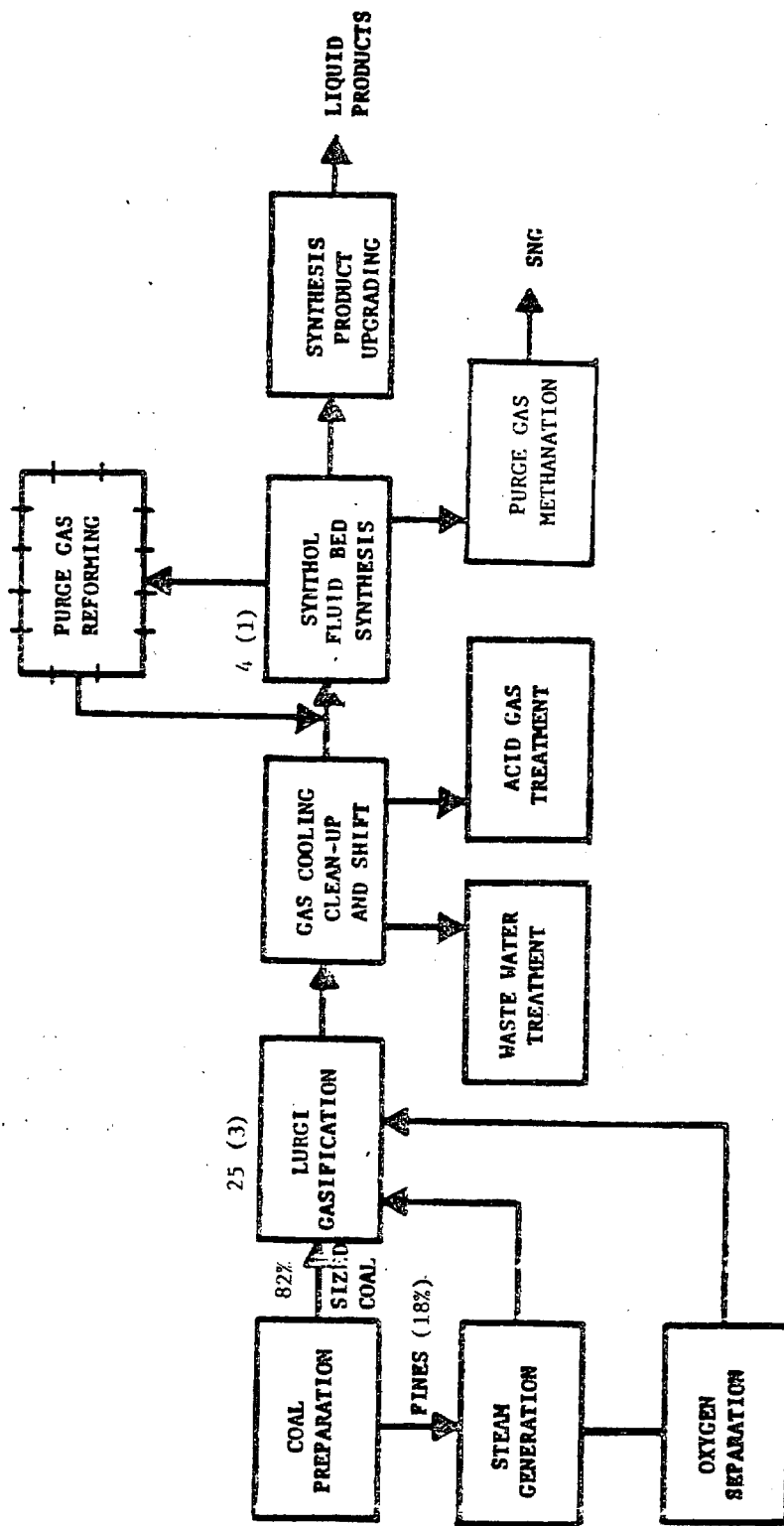
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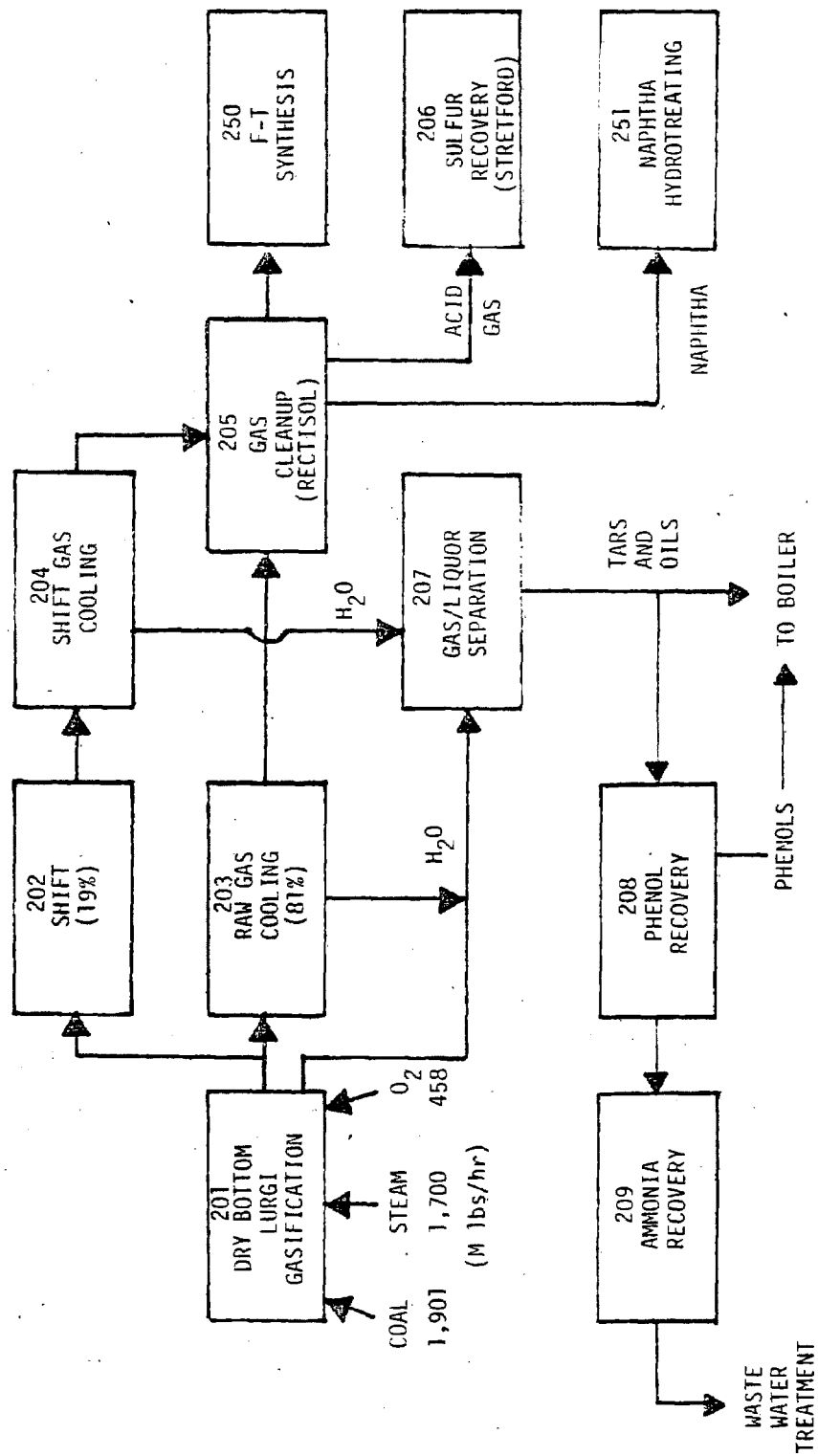FIGURE 4-1
MOBIL SASOL (U.S.) PROCESS FLOW

FIGURE 4-2

SASOL-U.S. SYNTHESIS GAS PREPARATION

Gasification is by fixed bed dry bottom Lurgi gasifiers. Twenty-eight gasifiers are available, with 25 required to produce the required quantity of synthesis gas. The remaining three are on standby or undergoing scheduled maintenance.

The Lurgi gasifier, described more fully in Section 5.0 below, is characterized by a relatively low temperature, low throughput, and high steam requirement. The gases produced are extremely wet and a high percentage (about 19 percent of the carbon content is in the form of methane. The H_2/CO ratio of the gas is approximately 2.0. After being subjected to a water quench to remove tars and particulates, the synthesis gas is split into two streams. About 20 percent enters a shift reactor where CO is reacted with water already present to produce additional H_2 and CO_2. The H_2/CO ratio of the shifted stream is approximately 10:1, which results in an overall H_2/CO ratio of about 2.5:1 when the shifted and raw streams are combined after being separately cooled.

Table IV-2 shows the characteristics of the gas after the shifted and unshifted streams are combined. Note that only about 2.9 M lb moles per hour of CO is shifted; thus only 2.9 M of 79.1 M lb moles of H_2O in the gasifier exit is actually consumed in the shift reaction. The balance is combined when the streams are cooled. Heat recovered during the cooling operation is used to produce low pressure steam in waste heat boilers.

TABLE IV-2

MATERIAL AND ENERGY FLUX IN SYNTHESIS GAS PREPARATION UNITS

(SASOL-U.S.)

	Dry Bottom Lurgi 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			33.7	1,484	36.6	1,610	2.9	128
CO			21.5	601	18.6	521	18.5	518
H_2			44.3	89	47.2	94	47.0	94
CH_4			12.9	206	12.9	206	12.5	200
C_2H_4/C_2H_6			.7	21	.7	21	.3	9
H_2O			79.1	1,424	.2	6	nil	nil
Misc. Gases			.5	12	.5	12	.3	8
Naphtha				16		16		
Tars/Oils/Phenols				110*				
				97				
Coal/(Ash)		1,901						
Steam		1,700						
Oxygen		458						
Totals		4,059	192.7	4,059	116.5	2,486	81.4	958
Energy (LHV) (MMBtu/hr)	15,561 (DAF Coal)		14,000		12,302		11,621	

*Includes Tars and Phenols recovered from quench water

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The gas is cleaned by a Rectisol process employing cold methanol absorption. The products of cleanup are a mixture of a clean synthesis gas and methane, an acid gas consisting of CO_2, sulfur and impurities, and gasifier naphtha. Synthesis gas and methane flow to the F-T synthesis reactors. Acid gases are sent to Stretford units for sulfur recovery. The gases are unsuitable for a Claus unit because of the very low H_2S concentration. Gasifier naphtha is hydrotreated and added to the gasoline pool.

4.2.3 Synthesis and Upgrading

Synthesis and upgrading in the SASOL-U.S. plant is shown schematically in Figure 4-3.

Synthesis is by Synthol fast fluid bed units developed by Kellogg and SASOL and successfully used in SASOL I and II. Four operating and one standby unit are employed. The units are approximately a 2:1 scale up of units operating at SASOL II.

The raw F-T products require upgrading or refining before they are able to meet product specifications for pipeline gas, oils and transportation fuels. The octane number of the raw gasoline is 55 (RON).[3] Seventeen processing units are used to produce finished products, 14 of them are used in the liquid product stream and three are used exclusively for pipeline gas manufacture. Table IV-3 lists the upgrading units and briefly describes their function in the overall scheme.
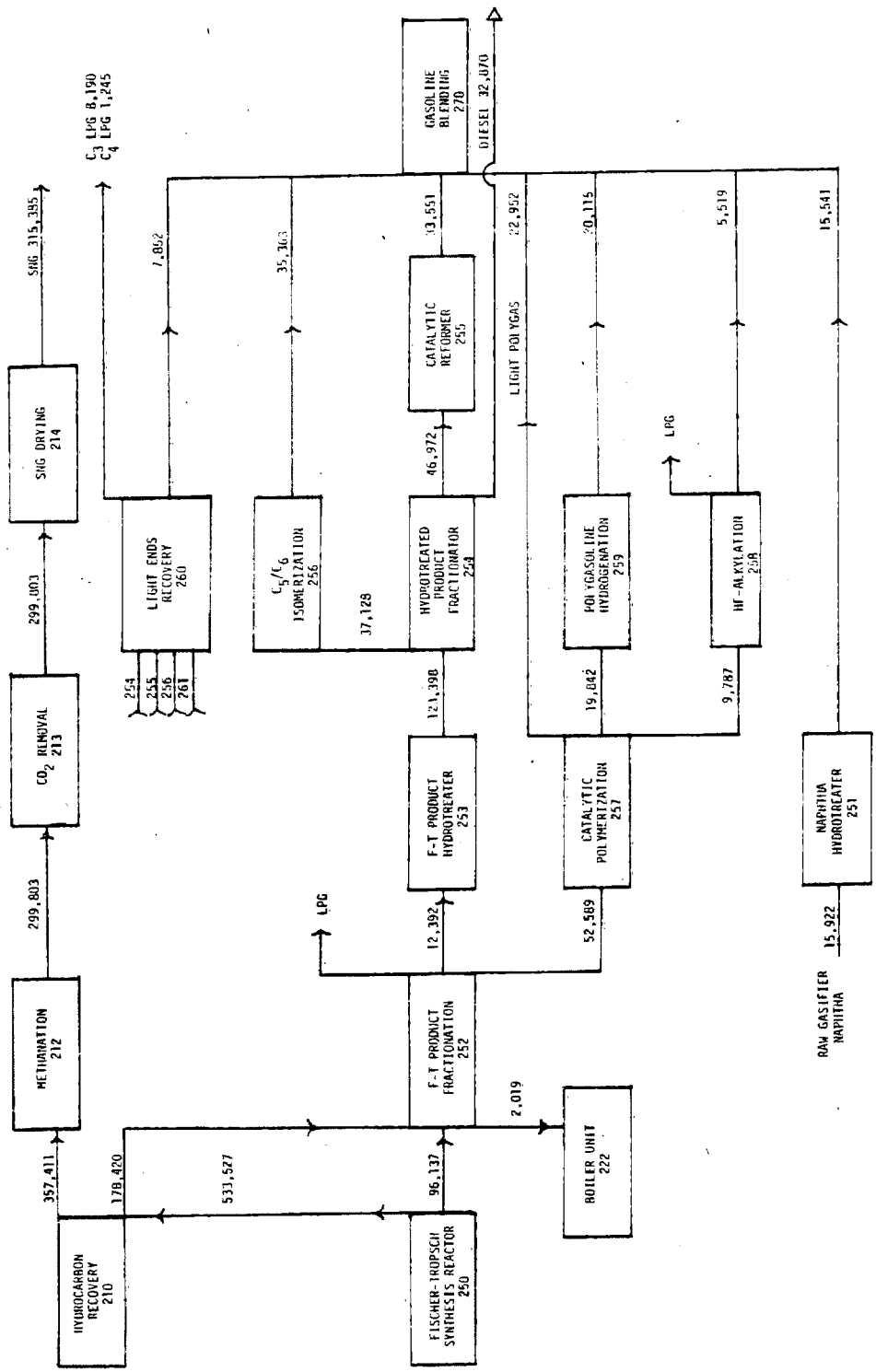
FIGURE 4-3
F-T DOWNSTREAM PROCESSING (SASOL-U.S.)
(UNITS LBS/HR)

TABLE IV-1

PRODUCT UPGRADING UNITS FOLLOWING F-T SYNTHESIS

No.	Unit	Process	Purpose
210	Hydrocarbon Recovery	Low Temp. Heptane Wash (prevents freezing out of CO_2)	Remove and recover heavy hydrocarbons from F-T syntheses purge gas to meet methanation feedstock specifications
211	Hydrogen Recovery	Pressure Swing Process (Union Carbide)	Recover hydrogen for use in the refinery
212	Methanation	Lurgi Process	SNG production (CO < 0.1%)
213	Carbon Dioxide Removal	Amine Guard Process (uses monoethanol amine MEA)	Lowers CO_2 content in SNG to 0.5%
214	SNG Drying/Compression	TEG Wash Process (Triethylene Glycol)	Dry SNG to 4 lb. moisture/MM SCF
251	Naphtha Hydrotreating	–	To produce clean stable naphtha for gasoline blending from raw gasifier naphtha
252	F-T Product Fractionation	4-Tower Dry System	Separates F-T products into light gases, feeds for polymerization and hydrogenation and 850^+ residuum for boiler fuel
253	F-T Product Hydrotreating	–	Saturates olefins and destroys remaining alcohols and acids
254	Hydrotreater Product Fractionation	Conventional 3-Tower System	Separates hydrotreated products into propane/hexane stream for isomerization, C_7 – 380° for reforming, diesel and heavy fuel oil products
255	Catalytic Reforming	Platformer	To increase anti-knock quality of C_7 – 380° naphtha
256	C_5/C_6 Isomerization	Pt. Cat. Promoted with Organic Chlorides	Increase anti-knock of pentanes and hexanes
257	Catalytic Polymerization	Catalysed by Acids 85% Conversion Exothermic Phosphoric on Kieselguhr (UOP)	Polymerize propene/butene to gasoline (dimers with butene/trimers with propene)
258	HF Alkylation	HF or H_2SO_4 Catalysed Highly Exothermic (Phillips/UOP/Stretford)	Increase gasoline yield by catalytic alkylation of iso-butane and unpolymerized C_3 and C_4 olefins (also n-butane and propane LPG produced)
259	Poly-gasoline Hydrogenation	–	Saturates olefins in heavy cat. poly. gasoline
260	Light Ends Recovery	–	Separates light gasoline and iso-butane from hydrocarbons rich off gases
261	Hydrogen Purification	Commercial Cryogenic Technology (Linde)	Obtain hydrogen of 90% purity from hydrogen rich off gases
262	Alcohol Recovery	–	Prepare marketable alcohol mixture and recover methanol for Rectisol

For pipeline gas the main unit is the methanator which converts residual CO and some of the CO_2 into methane for pipeline gas production. Primary gasifier methane passes through the unit unchanged. Other reactions in the unit saturate ethylene and crack the ethane to methane. The other two units in the pipeline gas preparation train lower the CO_2 and water contents to SNG specifications.

The dry-bottom Lurgi gasifier produces a primary gasifier naphtha which must be hydrotreated before being blended with the upgraded naphthas from isomerization, alkylation and reforming. The naphtha hydrotreator performs this function by removing sulfur, nitrogen and oxygen feedstock impurities by hydrotreating the raw feed over a cobalt molybdenum catalyst. Hydrogen requirements for this process are met by the hydrogen recovery unit in the F-T purge gas stream.

The F-T product fractionator splits the main raw product stream into four streams. The light ends (mostly C_2) are sent to the pipeline gas train. The bulk of the liquid is sent to be hydrotreated. Reactive C_3 and C_4 olefins are sent to be catalytically polymerized and heavy fuel oils and residuals are sent to the boiler unit for firing.

Catalytic polymerization is an important process for refining F-T products because of the high concentration of C_3 and C_4 olefins produced. This polymerization is catalyzed by acids and is generally

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an 85% conversion process yielding trimers with propene feeds and dimers with butenes. These polymerized products are blended into gasoline.

Iso-butane and C_3/C_4 olefins are fed to an HF catalyzed alkylation unit where the resulting alkylate is also blended with the other gasoline fractions.

After the main liquid product has been hydrotreated, it is fractionated into three streams. The light fraction containing C_5 and C_6 components is sent to a platimum catalyzed isomerization unit where the pentanes and hexanes are isomerized to increase the anti-knock quality of the gasoline. The C_7^+ fraction is sent to a catalytic reforming unit and the heavy ends are fractionated into diesel, fuel oil and heavy fuel oil. The various gasoline producing streams are blended toghether to form the final products. The overall processing scheme is designed to maximize the quality and quantity of motor gasoline. The intermediate octane of the finished gasoline $\frac{R + M^*}{2}$ is 87. The diesel, fuel oils, alcohols, LPG and SNG, which are co-products of the refinery, are also of excellent quality and designated to meet current commerical product quality.

4.2.4 Auxiliary Processes

Auxiliary processes which are part of the SASOL gasification section are listed below:

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*R&M refer to research octane and motor octane numbers

UNIT	PROCESS
206	Sulfur Recovery
207	Gas/Liquor Separation
208	Phenol Recovery
209	Ammonia Recovery

The sulfur recovery unit receives the acid gas removed by the gas purification system. A Stretford process is used to recover elemental sulfur. The cooled liquor from the gasifier and gas coolers flows to a gas/liquor separation unit for recovery of tars and oils. Phenols are recovered from the liquor by a Phenolsolvan process using di-isopropyl ether as a solvent. Ammonia is then recovered by a Chemie Linz/Lurgi process employing stripping and scrubbing techniques without the addition of chemicals or solvents.

4.2.5 Offsites

Offsite units which are essential to SASOL are listed below. All offsite units employ commercially available state-of-the-art processes.

Offsite Units

221	Oxygen Production
222, 223	Boiler, Main Superheater
225 & 226	Stackgas Precipitator and Clean-Up
224	Superheater
227	Instrument and Plant Air
228	Coal Handling
229	Ash Handling
231	BFW Preparation
232	CW Make-Up Preparation
233	CW Towers
234	Power Generation
235	Waste Water Treatment
236	Blow-Down Facilities
237	Storage

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238	Interconnecting Piping
241	Refrigeration
270	Gasoline Blending
271	F-T Catalyst Preparation
	Miscellaneous (Water Ponds, Misc. Tankage)

The most costly offsites are Unit 221, Oxygen Plant, and Units 222 through 226 which comprise the steam plant. Oxygen production is by air separation using steam driven compressors. 14,060 lb moles/hr of 98 percent purity oxygen are produced.

The steam plant produces 3,111 M lb/hr of HP steam for the oxygen plant, the Lurgi gasifier and other plant needs. In addition 856 M lb/hr of MP and LP steam produced in waste heat boilers is superheated. Fuel for the steam plant consists of 415 M lb/hr of coal fines plus 107 M lb/hr of tars, oil and phenols recovered from the gasifier output. These supplemental fuels account for 32 percent of the steam plant fuel. SASOL II recovers these supplemental fuels as by-products, which is the major reason why the percent of total coal used for steam generation at SASOL II is higher than in the SASOL-U.S. plant described here. However, steam coal requirements at SASOL II are further compounded by the poor quality of the SASOL coal. It has a 12 percent lower HHV than the Western coal assumed herein, but is less reactive, and thus requires more steam and oxygen for gasification.

4.3 Material and Energy Balance

The overall material balance for the SASOL-U.S. plant is shown in Table IV-4. The energy content of each stream, and the resulting

TABLE IV-4

MATERIAL BALANCE[1]

INPUT

Coal, MST/SD	27.8
Water, gpm	6,600

PRODUCT

Gasoline, B/Standard day (SD)	13,580
SNG, MMSCF/SD	173.3
C_3 LPG, B/SD	1,107
Butanes, B/SD	146
Diesel Fuel, B/SD	2,307
Fuel Oil, B/SD	622
Alcohol, Mlb/SD	510
Methanol, B/SD	—
Total Liquids B/SD	19,591
Fuel Oil Equivalent, B/SD[*]	44,950

OTHER

Power, MW(e)	3.31
Coal Fines, MST/SD	0
Ammonia and Sulfur[**]	

EFFICIENCY (HHV), % (TOTAL)	58
EFFICIENCY (HHV), % (FUELS ONLY)	57

YIELDS PER TON OF DRY COAL (B)

Liquid Fuels	.98
Liquid Fuels C_4+	.92

[*]All products converted to FOE barrels @ 6.0
   MM Btu/B.

[**]Yields for each case: NH_3 = 103 ST/SD;
   S = 61 ST/SD

thermal efficiency of the plant is given in Table IV-5. Figure 4-4 shows the energy losses in each section of the SASOL-U.S. plant which result in the final efficiency of 57% (HHV). The figures at the top of the chart show the energy flux in MM-Btu/hr and the efficiency of each major section of the plant. Unit efficiencies are shown in parentheses on applicable units.

The largest loss is seen to occur in the gasification section in which the overall efficiency is 70.5%. The major losses result from the diversion of coal, tar oils and phenols for production of steam. If computed on the basis of coal gasified rather than total coal, the HHV efficiency of the gasifier would be 86.7 percent.

Gas cleaning and shift losses are seen to be 4.1%. The shift reduces the steam HHV by less than 1%. The major loss occurs from the removal of methane, ethane and ethylene in the Rectisol unit. About 2/3 of the ethane and ethylene and about 5% of the methane are lost. These lost fuels account for over 3% of the energy in the gas stream. The lost fuels are incinerated in the boiler along with acid gases remaining after sulfur recovery, but no energy recovery is credited.

The synthesis section is seen to have an overall efficiency of 86.8 percent. The efficiency of the Synthol synthesis unit (excluding unreacted products) is 79%, which compares favorably with a theoretical efficiency of 81.1 percent. However, only about 1/2 of the syngas stream is reacted in the Synthol reactor, the balance being unreacted H_2 and CO, and light gases which pass through

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TABLE IV-5

SASOL-U.S. HEAT BALANCE[1]

	Quantity, Unit/SD	High Heating Value (HHV)	Total HHV, MMBtu/hr	Percent of Input	Low Heating Value (LHV)	Total LHV MMBtu/hr	Percent of Input
Input							
Coal, DAF	18,593 T	12,720Btu/lb	19,708	–	12,236Btu/lb	18,959*	–
Methanol	4 T	9,724Btu/lb	3	–	8,610Btu/lb	3	–
Total			19,711			18,962	
Output							
SNG	173.26MMSCF	1,003.3Btu/SCF	7,243	36.8	904.7Btu/SCF	6,531	34.4
C$_3$ LPG	1,107Bbl	3.815MMBtu/BBl	176	0.9	3.514MMBtu/Bbl	162	0.9
C$_4$ LPG	146Bbl	4.321MMBtu/Bbl	26	0.1	3.991MMBtu/Bbl	24	0.1
10 RVP Gasoline	13,580Bbl	5.023MMBtu/Bbl	2,842	14.4	4.682MMBtu/Bbl	2,649	14.0
Diesel Fuel	2,307Bbl	5.345MMBtu/Bbl	514	2.6	4.994MMBtu/Bbl	480	2.5
Hvy. Fuel Oil	622Bbl	5.682MMBtu/Bbl	147	0.7	5.326MMBtu/Bbl	138	0.7
Subtotal			10,948	55.5		9,984	52.6
Alcohols	510Mlb	13,630Btu/lb	290	1.5	12,380Btu/lb	263	1.4
Sulfur	61 T	3,780Btu/lb	19	0.1	3,780Btu/lb	19**	0.1
Ammonia	103 T	9,693Btu/lb	83	0.4	8,020Btu/lb	69	0.4
Power	3.31 MW(e)	3,415Btu/kWH	11(2)	0.1	3,415Btu/kWH	11	0.1
Total			11,351	57.6		10,346	54.6

*If LHV of as mined coal used (7,893 Btu/lb), overall LHV thermal efficiency becomes 56.6%.
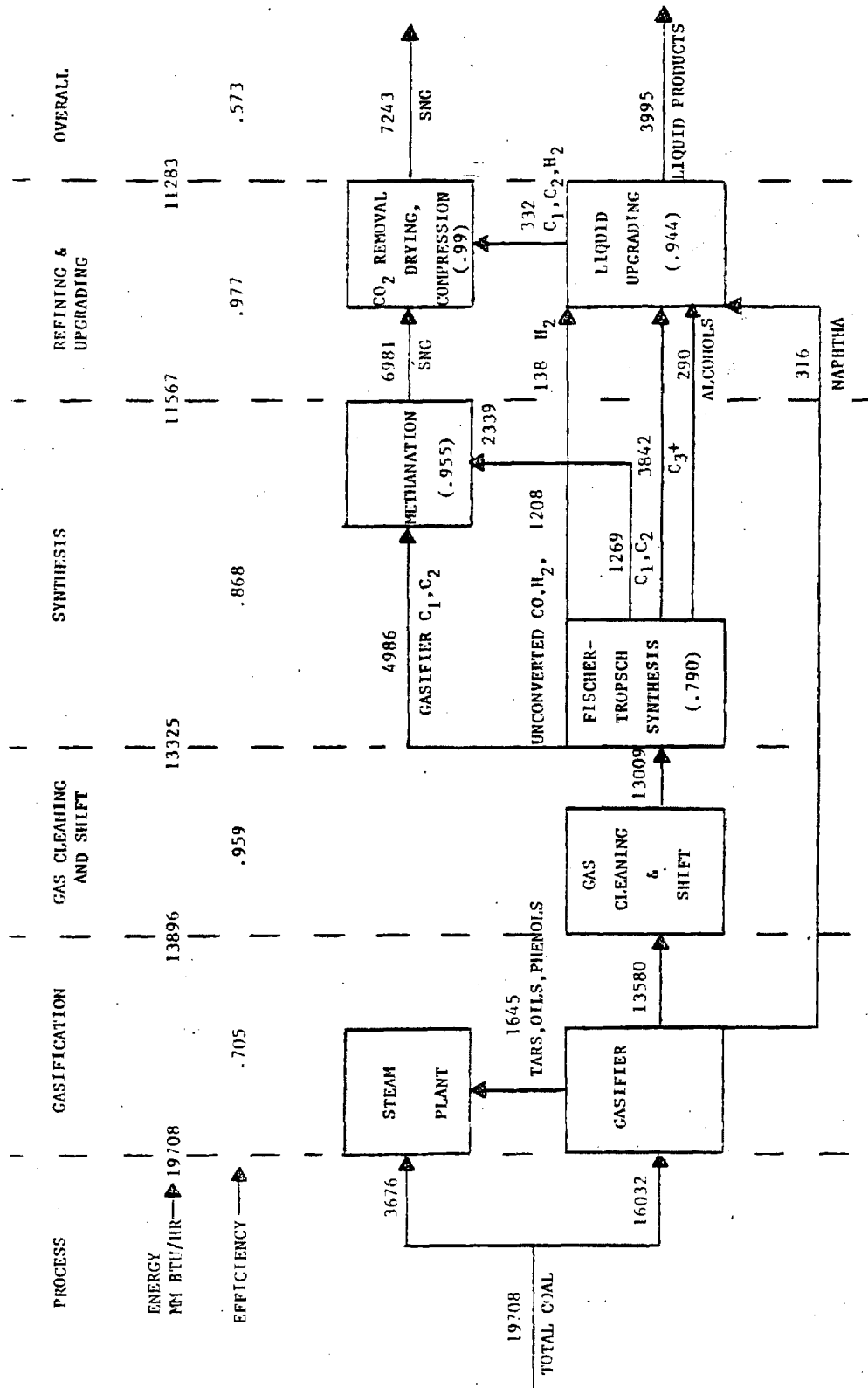
**Direct thermal conversion used.

FIGURE 4-4
SYSTEM ENERGETICS FOR SASOL-U.S. (MIXED OUTPUT)
(UNITS MM BTU/HR)

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unchanged. These gases, plus hydrocarbons lighter than C_3 which are formed in the synthesis unit enter the methanator which converts CO, H_2 and some C_2 to methane. The 95.5 percent efficiency of the methanator is computed based on total throughput of the units, most of which enters and leaves as methane. The conversion efficiency for products actually converted probably approaches the theoretical value of 85%.

Overall refining and upgrading efficiency, based on throughput, is 97.7 percent. This high value results because nearly 2/3 of the product is SNG which is refined and upgraded with a nominal (assumed 1%) loss. The efficiency of the liquid upgrading process is seen to be 94.4 percent. This value is based on thermal content of entry and exit streams, and does not account for the thermal value of steam supplied from other processes and/or the steam plant.

It should be noted that only 3,996 MM Btu/hr of the plant output is in the form of liquid fuel, while the bulk of the output (7,243 MM Btu/hr) is SNG. The relatively high efficiency of the overall plant results from the fact that most of the SNG consists of methane produced in the Lurgi gasifier. Liquids formed by F-T synthesis are subjected to synthesis losses (~20%) and upgrading losses (about 6%) before becoming marketable products.

Details of the yield structure from the SASOL Synthol reactors were taken from Reference 1, the MRDC Report. DOE has been informed by SASOL and FLUOR that a SASOL plant built for the U.S. market

would employ SYNTHOL units and upgrading refinements which would improve the overall ratio of liquid to SNG output. Details of these refinements are proprietary and would not be used in this study.

4.4 Capital and Operating Costs

The SASOL-U.S. plant designed by MRDC is estimated to have a construction cost of $1,186.1 million in October 1977 dollars. The breakdown of cost by operating unit is shown in Figure 4-5.

It is evident from Figure 4-5 that the bulk of costs is associated with producing synthesis gases. Only 222.8 M or 19% of the plant construction costs are for Fischer-Tropsch synthesis and the upgrading of synthesis products.

Table IV-6 shows the total plant investment breakdown, including engineering and design, contractor fees, other project costs, expense capital and working capital. The total, including a 15 percent estimating allowance, is seen to be $1,946.5 million or 1.59 greater than the estimated construction cost shown in Figure 4-6.

Annual operating cost estimates are shown graphically in Figure 4-6.

Coal costs, estimated at $7 per ton, account for $65.3 MM or 30% of the estimated annual operating cost of $216.8 MM. Maintenance, local taxes and insurance, and administration and overhead, in that order, are the remaining cost elements of significance.
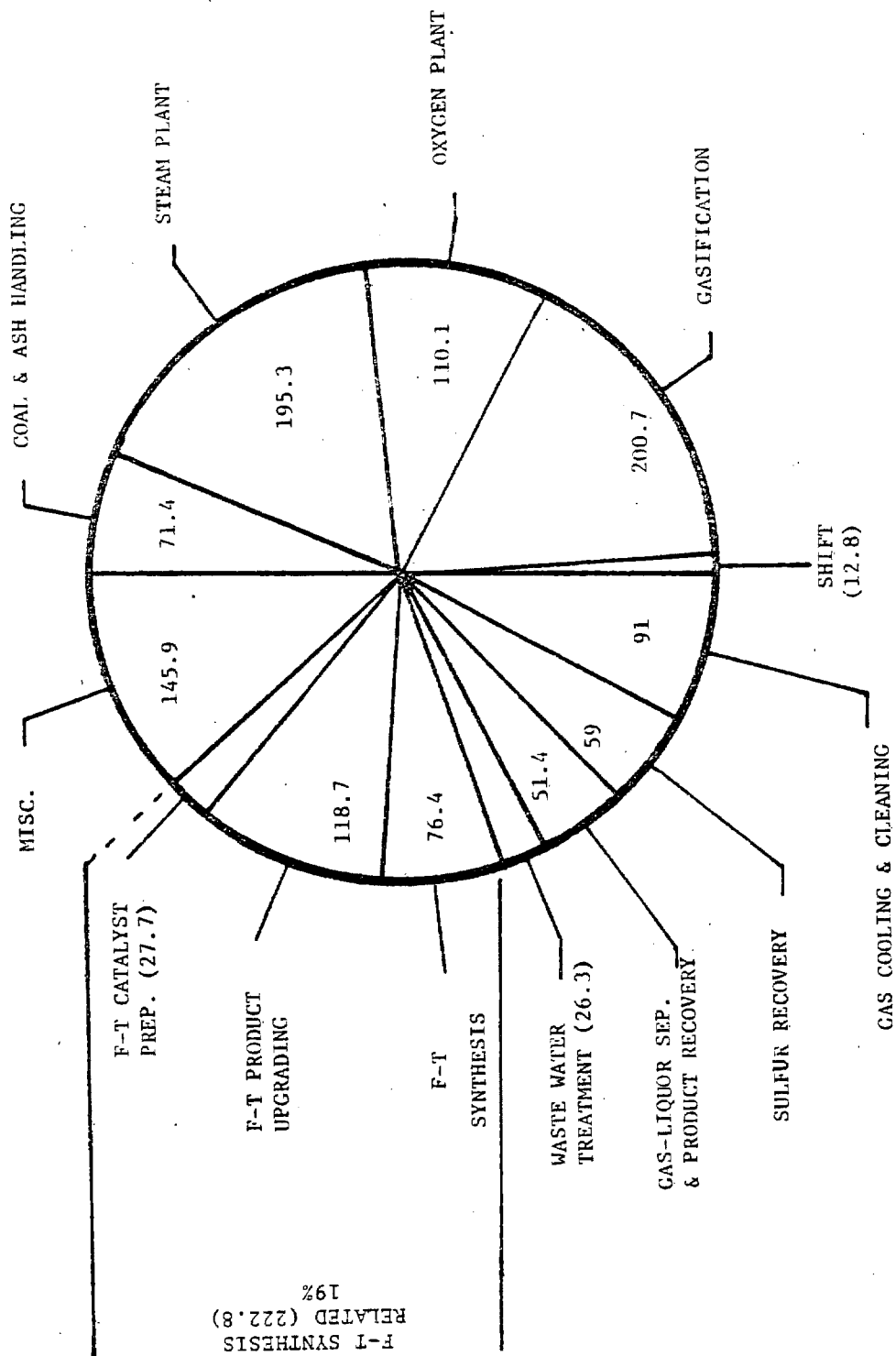
OXYGEN PLANT

STEAM PLANT

COAL & ASH HANDLING

GASIFICATION

110.1

195.3

200.7

71.4

SHIFT
(12.8)

91

145.9

59

51.4

MISC.

118.7

76.4

F-T CATALYST
PREP. (27.7)

F-T PRODUCT
UPGRADING

F-T
SYNTHESIS

WASTE WATER
TREATMENT (26.3)

GAS-LIQUOR SEP.
& PRODUCT RECOVERY

SULFUR RECOVERY

GAS COOLING & CLEANING

FIGURE 4-5
SASOL-U.S. PLANT CONSTRUCTION COSTS
TOTAL = $1186.1 M (1977)

F-T SYNTHESIS
RELATED (222.8)
19%

TABLE IV-6

SUMMARY OF CAPITAL COSTS[1]
($M 1977)

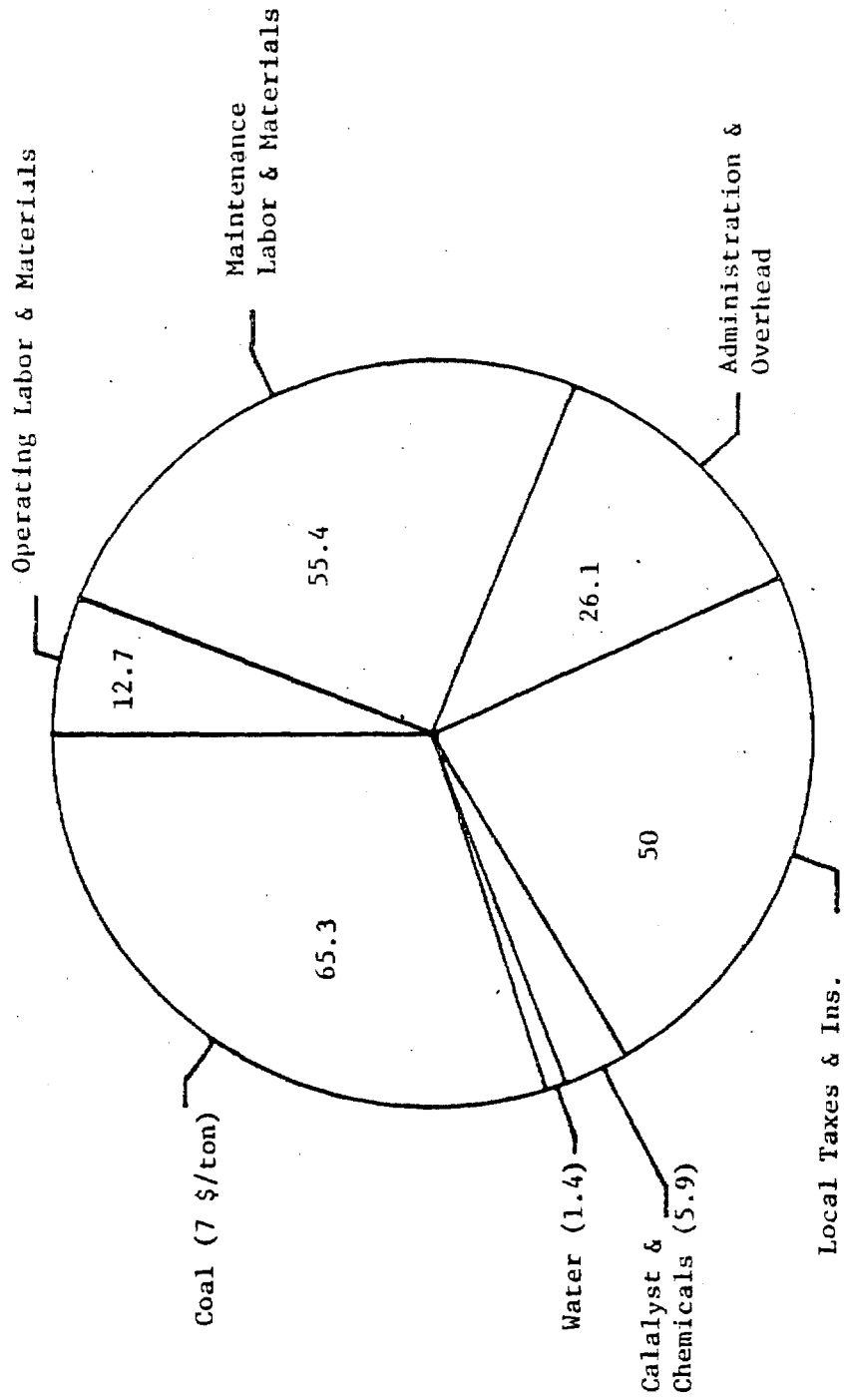|	
Construction Cost	1186
Engineering & Design	136
Contractor Fee	16
Overtime Premium	147
Workers Comp	80
Project Management	33
Other	46
Estimating Allowance (15%)	242
Total Capital Cost	1886

Capital Cost = 1.59 x Construction Cost

FIGURE 4-6
SASOL-U.S. OPERATING COST
TOTAL = $216.8 M/YR (1977)

Operating Labor & Materials

Maintenance
Labor & Materials

Administration &
Overhead

12.7

55.4

26.1

65.3

50

Coal (7 $/ton)

Water (1.4)

Calalyst &
Chemicals (5.9)

Local Taxes & Ins.

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