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RESEARCH GUIDANCE STUDIES TO ASSESS GASOLINE FROM COAL BY
METHANOL-TO-GASOLINE AND SASOL-TYPE FISCHER-TROPSCH
TECHNOLOGIES

Final Report

By
Max Schreiner

MASTER

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MAX SCHREINER

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ABSTRACT

This study provides a technical and economic comparison between the new Mobil methanol-to-gasoline technology under development and the commercially available Fischer-Tropsch technology for the production of motor gasoline meeting U.S. quality standards. Conceptual plant complexes, sited in Wyoming, are complete grass-roots facilities. The Lurgi dry-ash, pressure technology is used to gasify sub-bituminous strip coal. Except for the Mobil process, processes used are commercially available. Coproduction of products, namely SNG, LPG and gasoline, is practiced. Four sensitivity cases have also been developed in less detail from the two base cases.

In all areas, the Mobil technology is superior to Fischer-Tropsch: process complexity, energy usage, thermal efficiency, gasoline selectivity, gasoline quality, investment and gasoline cost. Principal advantages of the Mobil process are its selective yield of excellent quality gasoline with minimum ancillary processing. Fischer-Tropsch not only yields a spectrum of products, but the production of a gasoline meeting U.S. specifications is difficult and complex. This superiority results in about a 25% reduction in the gasoline cost.

Sensitivity study conclusions include: (1) the conversion of methanol into gasoline over the Mobil catalyst is highly efficient, (2) if SNG is a valuable product, increased gasoline yield via the reforming of SNG is uneconomical, and (3) fluid-bed operation is somewhat superior to fixed-bed operation for the Mobil methanol conversion technology.

SUMMARY

The purpose of this study is to provide a technical and economic comparison between the new Mobil methanol-to-gasoline technology and the commercially available Fischer-Tropsch technology for the production of motor gasoline meeting U.S. quality standards. Four additional sensitivity cases have also been developed in less detail from the two base cases.

The conceptual plant complexes are self-supporting, grass roots facilities located in Wyoming. The feedstock is a high volatile, sub-bituminous strip coal. Plant size is equivalent to the proposed 280 MMSCFD SNG plants. Coal gasification is by commercial Lurgi dry ash, pressure gasifiers. Except for the Mobil methanol conversion technology, all processes employed are commercially available. Plant design recognizes current environmental standards. Coproduction of all products has been generally practiced with the Lurgi gasifier methane product being marketed as SNG. No marketing restrictions have been imposed. Sufficient processing steps, however, are included to provide liquid products meeting U.S. market specifications and to produce an SNG which is interchangeable with natural gas. The study bases adopted have been guided by, but do not duplicate, those established in "Coal Gasification Commercial Concepts Gas Cost Guidelines". (Reference 1) The study scope and design bases are discussed in Sections 1 and 2.

Case definitions are briefly summarized below:

BASE CASES

<u>CASE</u>	<u>I</u>	<u>II</u>
Technology	Mobil Methanol Conversion	Fischer-Tropsch
Reactor Type	Fixed-Bed	Fluid-Bed
Major Products		
SNG	X	X
LPG	X	X
Gasoline	X	X
Fuels, Alcohols, etc.	-	X

SENSITIVITY CASES

CASE	I-A	I-B	I-C	II-A
Technology	Lurgi Methanol	Mobil Methanol	Mobil Methanol	Mobil Direct Route
Reactor Type	Tubular	Fixed-Bed	Fluid-Bed	Fluid-Bed
Major Products				
SNG	X	-	X	X
LPG	-	X	X	X
Gasoline	-	X	X	X
Fuels, Alcohols, etc.	-	-	-	-
Methanol	X	-	-	-

As stipulated in the contract, a fifth sensitivity case, the substitution of an advanced gasifier for the commercial Lurgi, was to be included. During its development using the Winkler pressure gasifier, it became apparent that sensitizing a different gasifier type onto a Lurgi base case is not practical. A new base case is required. Consequently, this effort is not reported.

Out of necessity, an additional sensitivity case - production of SNG only - was developed in order to obtain the SNG prices, or values, which are used in the coproduction cases for the determination of the gasoline cost. This case is briefly described in Appendix D.

The material balances for the study cases are summarized in Table 1. The superior gasoline selectivity of the Mobil methanol conversion technology is readily apparent; the 10 RVP gasoline yield is greater by about 60%. The Fischer-Tropsch technology produces a spectrum of compounds resulting in diesel and fuel oils and mixed alcohols as finished products. In addition, the Mobil technology gives a higher liquid product/SNG thermal ratio, 47/53 versus 35/65.

There are nine processing steps, common to both technologies, required for coal gasification and synthesis gas production. Nine additional steps are required for the Mobil route to achieve marketable products. The Fischer-Tropsch route, on the other hand, requires eighteen processing steps. Process descriptions and flow diagrams are included in Sections 3 and 4, in which the two base cases are discussed in detail.

Although both base cases yield gasolines meeting the target specifications, the Mobil process gasoline is of better quality, e.g., in research octane, 93 vs. 91, and in olefin content, 11 vs. 20%. In addition, the hydrotreated Lurgi gasifier naphtha with its high aromatic content, very low olefin content and good octane is an essential blending component in the F-T gasoline. The Mobil process gasoline quality, however, is not dependent upon the presence of this component.

Sensitivity Case I-A shows that the conversion of methanol to gasoline over the Mobil catalyst is highly selective and efficient. The major problem is the utilization of the additional production of low grade heat. Reforming the Lurgi gasifier methane in Sensitivity Case I-B increases the gasoline yield, about 70%, but with a severe 15% loss in the thermal efficiency. Fluid-bed operation for the Mobil methanol conversion technology, Sensitivity Case I-C, shows slightly improved gasoline yield and thermal efficiency over fixed-bed operation (Base Case I). The Mobil direct route technology, Sensitivity Case II-A, is a significant improvement over the F-T process. Gasoline upgrading is simplified and its yield is up by about 30%. Product selectivity as projected, however, is still poorer than for the Mobil methanol conversion technology. The sensitivity cases are discussed in Section 5.

Economics have been developed on both equity and utility financing. Unit costs are reported on two bases: (1) thermal basis in which each product Btu supports the plant cost and (2) multiple products basis in which plant gate values are assigned to coproducts other than gasoline and then the gasoline cost is determined. Investments and other costs have been developed specifically for the remote Wyoming site. All costs are in October, 1977 dollars. Inflation for a realistic plant start-up, e.g., in 1990, is not included. Detailed discussions of the bases are presented in the Cost and Economic Sections 6 and 7.

The economic results are summarized in Table 2. These conceptual plants are expensive, costing about 41.5 M\$ per calendar FOE barrel of total product. The 155 M\$ investment reduction for the Mobil Base Case over the F-T Base Case is due largely to the fewer gasoline processing units and corresponding reduced utility requirements. The conversion of methanol into motor gasoline adds from 90 to 150 MM\$ to the plant investment, depending upon the reactor-type for the Mobil technology. Converting the SNG coproduct into gasoline adds about 250 MM\$ to the plant investment. The Mobil direct route technology has the potential of being no more expensive than the methanol route technology.

The gasoline cost comparison between the base case technologies is:

	<u>Gasoline Cost, ¢/gal</u>	
	<u>Study Basis</u>	<u>Gas Cost Guidelines Basis</u>
Mobil Gasoline	60 to 100	50 to 80
F-T Gasoline	70 to 135	-

The lower costs are based on utility-type financing and pricing all coproducts at the same thermal cost in \$/MMBtu. The higher costs are based on equity financing at 12% DCF and pricing the non-gasoline coproducts according to the scenario described in Section 7. Depending upon the pricing basis, gasoline via the Mobil technology is from 10 to 25% less expensive than gasoline via Fischer-Tropsch.

If the economic bases are those given in "Gas Cost Guidelines", gasoline unit costs are about 20% lower than those developed from the study bases. Because many studies reported in the literature do not use bases reflecting actual constructing and operating conditions in remote areas, or generally use optimistic bases, we recommend the usage of the 50 to 80 ¢/gal range in these inter-study comparisons.

If SNG is a valuable, marketable product, reforming the gasifier methane is not an economical method to increase the gasoline yield. The projected fluid-bed operation has an economic advantage, up to 7¢/gal, depending upon the SNG, C₃ LPG and excess butane pricing, over fixed-bed operation for the Mobil conversion technology. The Mobil direct route technology has the potential to be economically viable. Since a high H₂/CO ratio in the synthesis gas does not appear to be critical (unlike the commercial methanol synthesis and Fischer-Tropsch processes), this technology could be an excellent partner with advanced gasifiers which generally yield synthesis gases having low H₂/CO ratios.

Detailed breakdowns of the product costs are reported in Table 3.

The sensitivity of the gasoline cost to the principal economic study bases, e.g., project life, DCF rate, interest rate, income tax, etc., and to perturbations in the investment, coal price and operating cost is discussed in Section 7.

TABLE 1
MATERIAL BALANCES

CASE	I (BASE)	I-A	I-B	I-C	II (BASE)	II-A
<u>INPUT</u>						
Coal, MST/SD	27.3	27.3	27.4	27.3	27.8	27.3
Water, gpm	6,300	6,200	6,800	6,000	6,600	6,600
<u>PRODUCT</u>						
Gasoline, B/SD	22,045	1,315 ⁽²⁾	37,430	23,065	13,580	17,485
SNG, MMSCF/SD	148.5	146.4	0	152.6	173.3	177.9
C ₃ LPG, B/SD	1,555	-	2,665	1,790	1,107	1,675
Butanes, B/SD	2,205	-	3,800	0	146	0
Diesel Fuel, B/SD	-	-	-	-	2,307	-
Fuel Oil, B/SD	-	-	-	-	622	-
Alcohol, Mlb/SD	-	-	-	-	510	-
Methanol, B/SD	-	50,840	-	-	-	-
Fuel Oil Equivalent, B/SD(1)	45,550	47,880	36,200	45,560	44,950	45,650
<u>OTHER</u>						
Power, MW(e)	5.31	(6.55)	1.38	(0.08)	3.31	(1.01)
Coal Fines, MST/SD	1.2	1.6	0	1.6	0	0.3
Ammonia and Sulfur ⁽³⁾						
<u>EFFICIENCY (HHV), %</u>	62	66	47	63	58	60

(1) All products converted to FOE barrels @ 6.0 MM Btu/B.

(2) Gasifier Naphtha

(3) Yields for each case: $\text{NH}_3 = 103 \text{ ST/SD}$; $\text{S} = 61 \text{ ST/SD}$

TABLE 2
ECONOMICS

CASE	I (BASE)	I-A	I-B	I-C	II (BASE)	II-A
INVESTMENT, MM\$						
Onsite	762	682	874	728	826	746
Offsite	644	602	738	635	713	662
Other	326	303	374	317	348	319
TOTAL	1,732	1,587	1,986	1,680	1,887	1,727
WORKING CAPITAL, MM \$	58	50	76	57	60	52
COAL COST, MM\$/YR	64	64	64	64	65	64
OPERATING COST, MM\$/YR	132	122	149	128	152	134
BYPRODUCT CREDIT, MM\$/YR	8	9	6	9	6	6
EQUITY UNIT COST @ 12% DCF						
Thermal Product						
\$/MMBTU	6.99	6.15	9.98	6.80	7.78	6.99
Gasoline, ¢/gal(1)	85	40(2)	121	83	93	85
Multiple Products						
Gasoline, ¢/gal	98	40(2)	124	93	133	106
UTILITY UNIT COST						
Thermal Product						
\$/MMBTU	5.08	4.47	7.24	4.94	5.67	5.09
Gasoline, ¢/gal(1)	62	29(2)	88	60	68	62
Multiple Products						
Gasoline, ¢/gal	71	29(2)	90	67	94	76

(1) At 5.1 MMBTU/B (Cases I, I-B, I-C and II-A); 5.0 MMBTU/B (Case II)

(2) Methanol

TABLE 3
PRODUCT UNIT COST BREAKDOWN

CASE	I (BASE)	I-A	I-B	I-C	II(BASE)	II-A
EQUITY UNIT COST @ 12% DCF						
Thermal Product						
Gasoline \$/MM Btu	6.99	6.15	9.98	6.80	7.78	6.99
¢/gal (2)	85	40(3)	121	83	93	85
SNG \$/MM Btu	6.99	6.15	-	6.80	7.78	6.99
C ₃ LPG \$/MM Btu	6.99	-	9.98	6.80	7.78	6.99
Butanes \$/MM Btu	6.99	-	9.98	-	7.78	-
Diesel Oil \$/MM Btu	-	-	-	-	7.78	-
Hvy F.O. \$/MM Btu	-	-	-	-	7.78	-
Alcohols \$/MM Btu	-	-	-	-	7.78	-
Multiple Products (1)						
Gasoline \$/MM Btu	8.07	6.13	10.19	7.63	11.12	8.69
¢/gal (2)	98	40(3)	124	93	133	106
SNG \$/MM Btu	6.17	6.17	-	6.17	6.17	6.17
C ₃ LPG \$/MM Btu	6.17	-	6.17	6.17	6.17	6.17
Butanes \$/MM Btu	7.77	-	9.89	-	10.82	-
Diesel Oil \$/MM Btu	-	-	-	-	10.14	-
Hvy F.O. \$/MM Btu	-	-	-	-	9.22	-
Alcohols \$/MM Btu	-	-	-	-	11.00	-
UTILITY UNIT COST						
Thermal Product						
Gasoline \$/MM Btu	5.08	4.47	7.24	4.94	5.67	5.09
¢/gal(2)	62	29(3)	88	60	68	62
SNG \$/MM Btu	5.08	4.47	-	4.94	5.67	5.09
C ₃ LPG \$/MM Btu	5.08	-	7.24	4.94	5.67	5.09
Butanes \$/MM Btu	5.08	-	7.24	-	5.67	-
Diesel Oil \$/MM Btu	-	-	-	-	5.67	-
Hvy F.O. \$/MM Btu	-	-	-	-	5.67	-
Alcohols \$/MM Btu	-	-	-	-	5.67	-
Multiple Products (1)						
Gasoline \$/MM Btu	5.84	4.43	7.40	5.51	7.38	6.29
¢/gal (2)	71	29(3)	90	67	94	76
SNG \$/MM Btu	4.51	4.51	-	4.51	4.51	4.51
C ₃ LPG \$/MM Btu	4.51	-	4.51	4.51	4.51	4.51
Butanes \$/MM Btu	5.54	-	7.10	-	7.08	-
Diesel Oil \$/MM Btu	-	-	-	-	7.06	-
Hvy F.O. \$/MM Btu	-	-	-	-	6.33	-
Alcohols \$/MM Btu	-	-	-	-	11.00	-

- (1) Discussion in Paragraph 7.1.4 on page 210.
 (2) At 5.1 MM Btu/Bbl (Cases I, I-B, I-C & II-A)
 At 5.0 MM Btu/Bbl (Case II)
 (3) Methanol

SECTION 1 STUDY SCOPE

This study has been prepared as a joint effort by Mobil Research and Development Corporation, Lurgi Kohle und Mineraloeltechnik GmbH and American Lurgi Corporation.

1.1 BASE CASES

The primary purpose of this scoping study is to provide a technical and economic comparison between the commercial Fischer-Tropsch technology and the Mobil methanol-to-gasoline technology for the production of motor gasoline from a U.S. coal. The primary coal gasification step uses the Lurgi dry-ash, moving-bed pressure gasifier.

Base Case I is the co-production of gasoline and SNG using the Mobil methanol-to-gasoline technology currently being developed through joint funding by DOE and Mobil Research and Development Corp. The methanol is first produced from the coal derived synthesis gas.

Base Case II is the co-production of gasoline and SNG using the state-of-the-art Sasol-type Fischer-Tropsch technology. The F-T products undergo extensive upgrading to become marketable under U.S. specifications.

1.2 SENSITIVITY CASES

In addition to the above base cases, the following technical sensitivity cases have also been evaluated:

Case I-A uses the same technology and synthesis gas production as in Base Case I, except the Mobil methanol-to-gasoline technology is eliminated and methanol and SNG are the co-products.

Case I-B uses the same technology and synthesis gas production as in Base Case I except the co-production of SNG is eliminated. A reforming unit converts methane into synthesis gas which is recycled to produce additional methanol and, ultimately, additional gasoline.

Case I-C uses the same technology and synthesis gas production as in Base Case I except the Mobil methanol-to-gasoline process uses a fluid-bed reactor in place of the fixed-bed reactor used in Base Case I.

Case I-D uses the non-commercial Winkler fluid-bed pressure gasification technology in place of the commercial Lurgi gasification technology. The Winkler technology is under active development in Germany. No economics for this case, however, are reported. The Winkler gasifier, radically different from the Lurgi dry ash gasifier, requires its own base case. (See Sensitivity Section 5.)

Case II-A uses the same technology and synthesis gas production as in Base Case II except the Mobil direct route technology currently under development through joint funding by DOE and Mobil Research and Development Corp. is substituted for the Sasol-type Fischer-Tropsch.

1.3 PLANT COMPLEX DEFINITION

For each base case, a conceptual plant complex has been developed. This complex is a self-supporting, grass roots facility encompassing all process and offsite units. Except for the Mobil and Winkler technologies, all technologies used in the plant design are commercially available. Location has been assumed to be adjacent a strip coal mine in the State of Wyoming. Products manufactured are available at the plant gate for shipment to the market place.

1.4 DATA SOURCES

The process information contained in this scoping study is based on either published or licensor data. (See References.) Much of the coal gasification information was derived from the proposed commercial Western SNG plants. In fact, the conceptual plant complexes developed in this study are nearly identical to these plants up through the production of clean synthesis gas. The combining of the processes used to obtain marketable SNG and gasoline from the synthesis gas, for the most part, is based on the in-house experience of the study participants. Although Mobil is the developer of the methanol conversion technology, MRDC made special efforts to assure that an impartial philosophy was maintained by the study participants.

Coal gasification and subsequent synthesis gas purification, methanol synthesis and SNG methanation are commercially available processes of Lurgi Kohle und Mineraloeltechnik.

The information for the methanol conversion technology is derived directly from published results for the process development studies conducted by Mobil under DOE contract E(49-18)-1773 with joint DOE and Mobil funding.

The basic data for the Fischer-Tropsch technology are derived principally from published literature reference. Some MRDC estimation was required for Case II development. Sasol did

not participate in the study, nor could the Department of Energy supply F-T data as originally anticipated. Consequently, MRDC cannot ascertain that the F-T technology used in Base Case II is exactly the commercially demonstrated Sasol technology.

Winkler Pressure Gasifier information was supplied by Davy Powergas, Inc.

No process or design optimization back-up studies have been made. In addition, no specific laboratory work was done for this study. As needed, in-house correlations, data banks, engineering programs, cost estimating programs, etc., were used.

Investment estimates are principally derived from in-house data. Where obtainable, vendor quotes, however, have been used for the offsite units.

1.5 SCOPE OF WORK

The work performed for each of the two base cases includes:

- material balance
- utility balance
- plant thermal efficiency
- block flow diagram
- simplified process flow diagrams
- equipment lists with major equipment dimensions

Each process unit is identified and briefly described with pertinent design considerations indicated. For the hydro-carbon upgrading processes, a technology critique is also included. In addition, the design bases used for the methanol-to-gasoline and the Fischer-Tropsch units have been outlined in some detail.

For each sensitivity case, the work performed has been reduced to include only:

- overall simplified material balance
- approximate utility balance
- plant thermal efficiency
- simplified block flow diagram

Discussion is limited to the changes in the base case brought about by the sensitivity.

Scoping or budget quality investment estimates have been developed for the base cases. The investment for each process and offsite unit is identified. Estimates of the manpower, catalyst requirements and chemical usages are also provided. The investment estimates for the sensitivity cases, for the most part, have been factored from the base cases. New sensitivity unit investments, e.g., methane reformer, have been developed.

The plant complex economics have been calculated using both utility financing and private investor financing. Sensitivity of the unit cost towards coal price, SNG price, byproduct values and investment and operating cost shifts has also been developed for the base cases. An abbreviated economic analysis is presented for the sensitivity cases.

1.6 UNIT SUMMARY - BASE CASES

The plant complex is comprised of a series of inter-connected process units, utility and support facilities (offsite units) and general plant facilities (infrastructure) as defined below:

PROCESS UNITS

<u>Unit No.</u>	<u>Description</u>
<u>Base Case I</u>	
101	Gasification (Lurgi Coal Pressure Gasification)
102	Raw Gas Shift
103	Raw Gas Cooling
104	Shifted Gas Cooling
105	Gas Purification (Rectisol)
106	Sulphur Recovery (Stretford)
107	Gas Liquor Separation
108	Phenol Recovery (Phenosolvan)
109	Ammonia Recovery (Chemie Linz/Lurgi)
110	Methanol Synthesis (Lurgi Low Pressure Process)
111	H ₂ Recovery (Pressure Swing)
112	Methanation
113	CO ₂ Removal (MEA Wash)
114	SNG Drying (TEG Wash)
150	Methanol Conversion (Mobil)
151	Naphtha Hydrotreating
152	Fractionation
153	HF Alkylation
<u>Base Case II</u>	
201 to 209	Identical to Base Case I
210	Hydrocarbon Recovery (Heptane Wash)
211	H ₂ Recovery (Pressure Swing)
212	Methanation
213	CO ₂ Removal (MEA Wash)
214	SNG Drying (TEG Wash) and Compression
250	F-T Synthesis
251	Naphtha Hydrotreating
252	F-T Product Fractionation
253	F-T Product Hydrotreating
254	Hydrotreated Product Fractionation
255	Catalytic Reforming
256	C ₅ /C ₆ Isomerization
257	Catalytic Polymerization
258	HF Alkylation
259	Poly Gasoline Hydrogenation
260	Light Ends Recovery
261	H ₂ Purification
262	Alcohol Recovery

OFFSITE UNITS
(Base Cases I and II)

<u>Unit No.</u>	<u>Description</u>
121/221	Oxygen Production (Air Separation)
122/222	Boiler
123/223	Main Superheater
124/224	Superheater
125/225	Electrostatic Stackgas Precipitator
126/226	Stackgas Clean-Up
127/227	Instrument and Plant Air
128/228	Coal Handling
129/229	Ash Handling
131/231	BFW Preparation (Deaerator & Demineralizer)
132/232	CW Make-Up Preparation
133/233	CW Towers
134/234	Electric Power Generation
135/235	Waste Water Treatment (Biological)
136/236	Relief and Blow-Down Facilities
137/237	Storage
138/238	Interconnecting Piping
141/241	Refrigeration
154/270	Gasoline Blending
271	F-T Catalyst Preparation

INFRASTRUCTURE
(Base Case I and II)

Office Buildings
Cafeteria
Maintenance Shops
Warehouse
Laboratory
Fire Protection System
Electric Distribution System
Truck and Railroad Unloading/Loading Facilities
Sewers
Roads and Parking Lots
Fencing
General Lighting
Communication and Security
Site Preparation

SECTION 2 STUDY DESIGN BASES

Where applicable, the plant design bases adopted for this study are generally those established by C.F. Braun & Co. for DOE in "Coal Gasification Commercial Concepts Gas Cost Guidelines" (Reference 1). A discussion of the principal bases follows.

2.1 PLANT CAPACITY

For both base cases, the plant capacity is based upon the gasification of 1,272 Mlb/hr of DAF coal. This amount of coal produces 742 MMSCF/SD of dry, purified synthesis gas for feed to methanol synthesis (Base Case I) or feed to Fischer-Tropsch synthesis (Base Case II). For reference, this capacity is equivalent to the capacity of the large 280 MMSCF/SD commercial SNG plants under consideration for the United States and, thus, is similar to the Guidelines plant size.

In the sensitivity cases, the gasification rate and synthesis gas yield are the same as used in the base cases, except for Sensitivity Case I-D. In this sensitivity, the 10 RVP gasoline product is the same as the Base Case I.

Because of differing energy requirements, the total coal charge, however, is not the same among the various cases.

2.2 DESIGN PHILOSOPHY

The overall plant complex has been designed using multiple trains, spare equipment and intermediate storage to provide an overall onstream factor of 92 percent. (See Sub-Sections 3.7 and 4.7.)

2.3 FEEDSTOCKS

The only feedstocks to the plant complex, besides air, are mined coal and fresh water. In addition, small quantities of chemicals, solvents and catalysts are required.

2.3.1 Coal

Gasification and boiler-firing are based on a coal having the properties listed in Table 2.1. This coal is a low sulfur, Wyoming sub-bituminous coal. Its properties differ slightly from the base Montana sub-bituminous coal in the Guidelines. It is a prime feedstock for the Lurgi dry-ash, pressure gasifier.

TABLE 2.1
STUDY COAL PROPERTIES
(Wyoming Sub-Bituminous Coal)

	<u>As Received</u>	<u>Dry and Ash Free (DAF)</u>
<u>Proximate Analysis, wt. %</u>		
Moisture	28.0	--
Ash	5.1	--
Fixed Carbon	33.8	50.5
Volatile Matter	33.1	49.5
	<u>100.0</u>	<u>100.0</u>

<u>Ultimate Analysis, wt. %</u>		
C		74.45
H		5.10
O		19.25
N		0.75
S		0.45
		<u>100.00</u>

Calorific Value, Btu/lb

High Heating Value (HHV)	8,509	12,720
Low Heating Value (LHV)	7,893	12,236

	-----Atmosphere-----	
<u>Fusion Properties of Ash, °F</u>	<u>Oxidizing</u>	<u>Reducing</u>
Softening Point	2,335	2,335
Melting Point	2,360	2,360
Flow Point	2,440	2,430

<u>Mineral Analysis of Ash</u>	<u>Wt. %</u>
SiO ₂	19.0
Al ₂ O ₃	14.0
Fe ₂ O ₃	6.0
MgO	7.8
CaO	36.0
Na ₂ O	2.7
K ₂ O	0.2
SO ₃	7.0
BaO	0.2
TiO ₂	1.0
P ₂ O ₅	1.8
Undetermined	4.3
	<u>100.0</u>

2.3.2 Fresh Water

The quality of the untreated fresh water assumed is the Western location water quality established in the Guidelines. In summary, the total dissolved solids are 496 ppm and the total hardness is 232 ppm CaCO_3 . It has been assumed that the water is available at the plant gate.

2.4 CLIMATIC CONDITIONS

Essentially, the climatic conditions established in the Guidelines have been used. In summary, they are:

Atmospheric pressure, psia	12.3
Air temperature (average), $^{\circ}\text{F}$	50
Relative humidity (average), %	70
Summer wet bulb, $^{\circ}\text{F}$	66
Summer dry bulb, $^{\circ}\text{F}$	95

2.5 PLANT SITE CONDITIONS

No major site development, other than leveling and grubbing, has been assumed. No piling is assumed to be required.

2.6 ENVIRONMENTAL REGULATIONS

The plant complex has been designed to meet all existing local and federal environmental regulations, as of July, 1977, for liquid and gaseous effluents.

Boiler stack gas clean-up facilities for SO_2 removal are provided to reduce the SO_2 emissions to 0.2 pounds of SO_2 per million Btu of fired heat. In addition, an electrostatic precipitator is included to reduce the particulate matter to 0.1 pound per million Btu of fired heat duty.

Although the sulfur emission target is more severe than the limitation in the Gas Cost Guidelines, this specification is the design limit for the latest proposed commercial Wyoming SNG plant (Panhandle Project). (Note: When calculating the "boiler" fired duty, the main superheater duty is included and LHV is the basis.)

Sulfur emissions from the process units are reduced by treating all sulfur containing offgases in a sulfur recovery unit. The tail gas stream containing 10 ppm H_2S from this unit is incinerated in the boiler to eliminate the hydrocarbons and to convert the COS and H_2S to SO_2 for removal in the stack gas clean-up facility. Only about 6% of the sulfur in the total coal charge escapes as SO_2 , amounting to about 5 T/SD of sulfur.

All aqueous liquid effluents are treated and recycled. Blowdown water is used for ash slurring, dust control, etc..

The wet ash and solid wastes, e.g., from stack gas clean-up and water treating, are sent to the mine for disposal.

2.7 PRODUCTS

Except as discussed below, this study is based on the concept that all products are marketable. Other than quality specifications, no marketing restrictions have been imposed. Products have been credited with their highest potential market value. The major products, however, are SNG and motor gasoline. A product summary is tabulated below:

<u>Product</u>	<u>Base Case</u>		<u>----Sensitivity Case----</u>				
	<u>I</u>	<u>II</u>	<u>I-A</u>	<u>I-B</u>	<u>I-C</u>	<u>I-D</u>	<u>II-A</u>
SNG	X	X	X	-	X	-	X
Methanol	-	-	X	-	-	-	-
Gasoline	X	X	-	X	X	X	X
Propane LPG	X	X	-	X	X	X	X
Butane LPG	X	X	-	X	-	X	X
Diesel Oil	-	X	-	-	-	-	-
Heavy Fuel Oil	-	X	-	-	-	-	-
Alcohols	-	X	-	-	-	-	-
Ammonia	X	X	X	X	X	-	X
Sulfur	X	X	X	X	X	X	X
Electric Power (excess)	X	X	X	X	X	X	X
Coal Fines (excess)	X	-	X	-	X	-	X

The Lurgi gasifier produces phenols, naphtha, oil and tar which are recovered during raw gas cooling. For expediency, the phenols, oil and tar are used as boiler fuel. Although technically possible, the upgrading of these materials into saleable products would have little, if any, effect on the plant complex economics. On the other hand, the gasifier naphtha is hydrotreated for gasoline blending, principally, because this aromatic stock is needed to yield a satisfactory 10 RVP gasoline in the Fischer-Tropsch case.

The Fischer-Tropsch reaction produces a small quantity of acids. They have been neutralized with caustic as recovery is not economical (8).

Ammonia and hydrogen sulfide obtained during raw syn gas purification are recovered as anhydrous ammonia and sulfur.

2.8 PRODUCT SPECIFICATIONS

2.8.1 SNG

The SNG product has been formulated to be interchangeable with pure methane as determined by three indexes. (See Guidelines.) The first two - lifting and flash-back - relate the dynamic equilibrium between the speed of the air-gas mixture issuing from the burner and the ignition velocity of the gas for the substitute gas to that for the standard gas. For example, a burner adjusted for pure methane will exhibit flame lifting if the inerts (N_2) concentration of a substitute gas is too high, or will have flash-back problems if the SNG concentration of hydrogen, with its high ignition velocity, is too high. The third - yellow-tip - relates the amount of olefins or illuminants in the substitute gas to the amount in the standard gas.

In addition, it meets the following limitations:

Carbon monoxide	0.1 vol. %
Hydrogen sulfide	0.25 grains/100 SCF
Total sulfur	10 grains/100 SCF
Water	4 lb/MMSCF,

and is available at the plant gate @ 1,000 psig.

2.8.2 Hydrocarbon Products

The target specifications are primarily based on the 1976 Annual Book of ASTM Standards and represent industry guidance for establishing product properties between buyer and seller. Properties and comments specific to the various hydrocarbon products can be found in the discussion and tables included in Sub-Section 3.2 and 4.2.

2.8.3 Alcohols

No specifications have been established. The alcohol product has been upgraded to be essentially free of water, acids, aldehydes and ketones.

2.8.4 Byproducts

Ammonia: Agriculture grade is produced. (With a relatively small increase in costs, chemical grade ammonia could be produced by the ammonia recovery process used in the study.)

Sulfur: The product is a liquid at 99.5% purity.

Electric Power: Excess electric power is produced at 6,000 volts.

2.9 COAL BALANCE

The number of Lurgi gasifiers is based on coal sized in the range of 1/4" to 2", containing not more than 7% undersize. The minus 1/4" coal is used as boiler fuel. When crushing the study coal, it has been estimated that the coal preparation plant yields 83.5% 2" to 1/4" gasifier coal and 16.5% minus 1/4" fines. Since the study basis calls for the combustion rather than the recovery of the gasifier tar, oil and phenols, a surplus of fine coal is possible. When this occurs, the excess fines are sold in accordance with the Guidelines.

2.10 UTILITY BALANCE

The general design bases calls for in-plant power and steam generation and, if possible, no export of these items. The utility balances were developed towards this goal, but detailed utility rebalancing calculations are outside the limits of this scoping study. Thus, a small surplus of electric power is available for export in the base cases, amounting to less than 5% of the total steam and electric power required to operate the complex. In the sensitivity cases, the balance swings from purchase to surplus electric power.

SECTION 3

BASE CASE I

MOBIL METHANOL-TO-GASOLINE TECHNOLOGY

BASE CASE I
3.1 MATERIAL BALANCE

The overall material balance is shown below:

<u>Input</u>	<u>Mlb/hr</u>
Coal, as mined	2,277.8(1)
Air	6,083.8
Water	3,143.7
	<u>11,505.3</u>

<u>Output</u>	<u>Mlb/hr</u>
Coal fines (excess)	102.5(2)
Ash	140.1
Products	530.9(3)
Byproducts	13.7
Blowdown water	809.5
Stack and evaporation losses	9,908.6
	<u>11,505.3</u>

(1) 27,334 T/SD

(2) 1,230 T/SD

(3) 45,550 FOE Bbl/SD @ 6 MM Btu/FOE Bbl

Overall plant consumptions per FOE barrel of product are:

Coal:	0.573 T
Water:	4.73 Bbl
Air:	1.60 T

Stack and evaporation losses amount to about 2.6 tons per FOE barrel of product. The barrels of methanol feed required to yield a barrel of 10 RVP gasoline (excluding the Lurgi gasifier naphtha) is 2.44.

Appendix A contains the complete, detailed material balance for Base Case I.

BASE CASE I
3.2 PRODUCT YIELDS AND QUALITY

Stream numbers are given below for reference to the material balance sheets in Appendix A and process diagrams in Sub-Section 3.4.

3.2.1 SNG (49)

<u>Quantity</u>	16,306.3 lb-mol/hr
	265.4 Mlb/hr
	148.5 MMSCF/SD

Composition (Only compounds greater than 0.1% are listed: see Appendix A for detailed composition.)

	<u>Percent</u>
Hydrogen	1.7
Methane	95.5
Ethane	0.2
Propane	0.1
Butane	0.1
Carbon dioxide	0.5
Inerts (N ₂ + Ar)	1.9
	100.0

Other

Heat of combustion (HHV)	980 Btu/SCF
Carbon monoxide (0.1% max.)	0.02%
Water	0.01%
Sulfur	None

Compatibility Indexes (versus pure methane)

<u>Index</u>	<u>Calculated</u>	<u>Preferable</u>	<u>Objectionable</u>
Lifting, I _l	1.03	under 1.0	above 1.06
Flash-back, I _f	1.02	under 1.15	above 1.2
Yellow-tip, I _y	1.05	above 1.0	under 0.8

The SNG product is of satisfactory quality and is compatible with pure methane. The inerts from the oxygen stream for gasification, however, give rise to a lifting index value between the preferable and objectionable values. From a processing standpoint, there is little that can be done to reduce this value. The selection of a more expensive oxygen plant yielding a 99+% pure product, however, would alleviate this problem. (The SNG product, on the other hand, has a lifting index value better than the preferable value when compared with the high methane natural gas standard cited in the A.G.A. Research Bulletin Number 36. This gas has a 5.0% inerts content.)

3.2.2 Gasoline (50)

Quantity

22,045 Bbl/SD
235.7 Mlb/hr

Blending

<u>Component</u>	<u>Wt. %</u>
Mixed butanes	2.4
Alkylate	3.2
Stabilized gasoline	87.8
Hydrotreated gasifier naphtha	6.6
	<u>100.0</u>

Properties

Estimated properties and a comparison with unleaded gasoline specifications are presented in Table 3.2.1. Table 3.2.2 gives a more detailed listing of the principal unleaded gasoline specifications.

The 10 RVP gasoline is of excellent quality and meets all specifications. Its octane rating is estimated to be about one above the specification number. Volatility and composition are similar to present-day gasolines. Although the content of the troublesome durene is below the target maximum of 5.0 wt. %, it would be preferable if it were below 4.0 wt. %.

Among the high molecular weight compounds in the product are the tetramethyl benzenes of which durene is one of three isomers. In view of durene's high melting point (174°F), and relatively high boiling point (386°F), a potential exists for durene crystallization at high concentrations under severe engine carburetor operating conditions. (Durene, an excellent octane material, is present only at low concentrations in conventional gasolines.) In chassis dynamometer testing of fuel mixtures, a 5 wt. % durene content frequently caused carburetor icing and engine stalling. (See Reference 2.)

3.2.3 Mixed Butanes (103)

Quantity

2,205 Bbl/SD
18.3 Mlb/hr

Properties

Estimated properties and a comparison with LPG specifications are shown in Table 3.2.3.

This product is a satisfactory commercial butane fuel. Because of the high iso-butane content of 76 vol. %, it also is a potential alkylation unit feedstock.

3.2.4 Propane LPG (102)

Quantity

1,555 Bbl/SD
11.5 Mlb/hr

Properties

Estimated properties and a comparison with LPG specifications are shown in Table 3.2.3.

This product is a satisfactory commercial propane fuel.

3.2.5 Byproducts

Sulfur (29)

Quantity
Impurities

61 T/SD
0.5 wt. %

Anhydrous Ammonia (17)

Quantity
Impurities (mainly water)
Grade

103 T/SD
0.5 wt. % max.
agricultural

Excess Power (6,000 volts)

5.31 MW(e)

3.2.6 Coal Fines (2.1)

Quantity

102.5 Mlb/hr

The thermal requirements do not require the use of all the fines produced in the coal preparation unit.

TABLE 3.2.1

COMPARISON OF PRINCIPAL UNLEADED GASOLINE SPECIFICATIONS
WITH ESTIMATED METHANOL-TO-GASOLINE GASOLINE PROPERTIES-
CASE I

	Estimated MTG Unleaded Gasoline Properties	Specification
Gravity, °API	61.4	
Octane Numbers		
Research	93	
Motor	83	82 min.
(Research + Motor)/2	88	87 min.
Volatility		
Reid Vapor Pressure, lb	10.0	
Distillation, °F		
IBP	85	
10%	110	158 max.
30%	146	
50%	200	170/250 min./max.
70%	262	
90%	336	374 max.
EP	388	437 max.
V/L Ratio (=20), °F	@ 129	@ 140 max.
Sulfur, wt. %	Nil	0.10 max.
Composition, vol. %		
Paraffins	51	
Olefins	11	20 max. (target)
Naphthenes	9	
Aromatics	29	
Durene Content, wt. %	4.6	5 max. (target)
Molecular Weight	94	

TABLE 3.2.2
PRINCIPAL UNLEADED GASOLINE SPECIFICATIONS

	ASTM D439 ⁽¹⁾	OTHER	POSSIBLE WAIVERS
Octane Numbers			
(Research + Motor)/2, min. (D2699 and D2700)	87		
Motor, min. (D2700)	82		
Volatility			
Reid Vapor Pressure, max. (D323)	15 ⁽²⁾		15.5-16.0 (in winter)
Distillation, °F (D-86)			
10% Evap. max. temp.	158 ⁽²⁾ (140 avg.)		
50% Evap. min./max. temp.	170/250 ⁽²⁾ (avg. 240 max.)		
90% Evap. max. temp.	374 ⁽²⁾ (370 avg.)		383 ⁽⁶⁾
End Point, max. temp.	437		446 ⁽⁶⁾
% Recovered, min.	-	95 ⁽³⁾	No Spec.
% Residue, max.	-	2 ⁽³⁾	
V/L Ratio, max. °F @ 20 (D2533)	140 ⁽²⁾		
Existent Gum, max. mg/100 mls (D381)	5		
Sulfur, max. wt. % (D1266)	0.10	0.05 ⁽⁴⁾	0.15 ⁽⁶⁾
Benzene, max. wt. %		5 ⁽⁷⁾	
Bromine Number, max. (D1159)	-	30 ⁽⁴⁾	
Induction Period, min. (D525)	-	240 ⁽⁵⁾	
Corrosion, max. (D130)	No. 1		
Lead Content, max. g/USG (D2547 or D2599)	0.05		

(1) 1976 Annual Book of ASTM Standards

(2) Lower max. limits apply according to season, geographical location, and individual state specification.

(3) Applicable in several states.

(4) California

(5) Recommended

(6) ASTM ES-1-74 (Emergency) specifications for conventional (not unleaded) gasoline.

(7) Federal specification.

TABLE 3.2.3

COMPARISON OF PRINCIPAL LPG SPECIFICATIONS WITH
ESTIMATED METHANOL-TO-GASOLINE LPG PROPERTIES-
CASE I

	<u>Commercial Propane ASTM D1835*</u>	<u>Est. MTG C₃ LPG Properties</u>	<u>Commercial Butane ASTM D1835*</u>	<u>Est. MTG C₄ LPG Properties</u>
-27- Vapor Pressure at 100°F, max. psig (D1267)	208	171	70	52
Volatile Residue				
Butane and Heavier, max. vol. %	2.5	1.9	-	-
Pentane and Heavier, max. vol. %	-	-	2.0	0.7
Corrosion, max. (D1838)	No. 1	No. 1	No. 1	No. 1
Sulfur, grains (D2784)	15	None	15	None

* 1976 Annual Book of ASTM Standards

BASE CASE I
3.3 THERMAL EFFICIENCY

Table 3.3.1 is a summary of the overall plant thermal efficiency, which is defined as the combustion heat of the products, byproducts and excess power divided by the combustion heat of the total coal feed. The efficiency has been calculated using both the high and low heating values. The HHV thermal efficiency is 62.2% and the LHV 59.2%.

The study basis calls for the burning of the oil and tar recovered from the raw synthesis gas in the coal-fired boiler. If these materials were severely hydrotreated into a quality fuel oil and replaced by coal fines, the overall plant efficiency increases by approximately 2% to 64.4%. The yield of fuel oil is about 5,700 B/SD; there is a small decrease in the SNG yield. The coal feed would increase to about 28,000 T/SD.

TABLE 3.3.1

METHANOL-TO-GASOLINE CASE I - THERMAL EFFICIENCY

	Quantity, Unit/SD	High Heating, Value (HHV)	Total HHV, MMBtu/hr	Percent of Input	Low Heating, Value (LHV)	Total LHV MMBtu/hr	Percent c Net Input
Input							
Coal, DAF	18,286 T	12,720 Btu/lb	19,383	-	12,236 Btu/lb	18,646 ⁽¹⁾	-
Fines(excess)	(823 T)	"	(872)	-	"	(840)	-
Net Coal	17,463		18,511			17,806	
Output							
SNG	148.52 MMSCF	980.4 Btu/SCF	6,067	32.8	882.6 Btu/SCF	5,462	30.7
C ₃ LPG	1,555 Bbl	3.816 MMBtu/Bbl	247	1.3	3.509 MMBtu/Bbl	227	1.3
C ₄ LPG	2,205 Bbl	4.191 MMBtu/Bbl	385	2.1	3.865 MMBtu/Bbl	355	2.0
10 RVP Gasoline	22,045 Bbl	5.105 MMBtu/Bbl	4,689	25.3	4.771 MMBtu/Bbl	4,382	24.6
Sub-Total	-	-	11,388	61.5	-	10,426	58.6
Sulfur	61 T	3,780 Btu/lb	19	0.1	3,780 Btu/lb	19	0.1
Ammonia	103 T	9,693 Btu/lb	83	0.5	8,020 Btu/lb	69	0.4
Power(excess)	5.31 MW(e)	3,415 Btu/kWH	18 ⁽²⁾	0.1	3,415 Btu/kWH	18 ⁽²⁾	0.1
Total			11,508	62.2		10,532	59.2

(1) If LHV of as mined coal used (7,893 Btu/lb), overall LHV thermal efficiency becomes 61.3%.

(2) Direct thermal conversion used.

BASE CASE I
3.4 PROCESS DESCRIPTIONS

The discussions below follow the block flow diagram (BFD) or the process flow diagram (PFD) as noted. These diagrams are placed at the end of the discussion. Detailed compositions and flows of the numbered streams are found on the material balance sheets in Appendix A.

3.4.1 General (BFD ZO-GEM-6894)

Sized coal is received from the offsite Coal Handling Unit 128 (see Sub-Section 3.5) and is gasified in Coal Pressure Gasification Unit 101. The raw gas generated is partially converted in Raw Gas Shift Unit 102 to meet Methanol Synthesis Unit 110 requirements, and is cooled in Raw Gas Cooling Unit 103 and Shifted Gas Cooling Unit 104 before flowing to Gas Purification Unit 105. This unit removes all sulfur compounds, CO₂ and naphtha from the raw gas. The purified gas (synthesis gas) is compressed and fed into Methanol Synthesis Unit 110. The crude methanol is used as feed to Methanol Conversion Unit 150, which converts the methanol into gasoline. Stabilized gasoline is recovered in Fractionation Unit 152. Additional gasoline is produced in HF Alkylation Unit 153 by alkylating the olefins recovered in Unit 152. C₃ LPG and mixed butanes are also recovered in the alkylation unit. The naphtha recovered within Gas Purification Unit 105 is upgraded in Naphtha Hydrotreating Unit 151.

The gasoline product is produced from stabilized gasoline, hydrotreated naphtha, alkylate and a portion of the mixed butanes in the offsite Gasoline Blending Unit 154. (See Sub-Section 3.5.)

The purge gas of Methanol Synthesis Unit 110, containing mainly the methane of the synthesis gas, is sent to the SNG upgrading train. This train consists of Methanation Unit 112, where unreacted CO and H₂ and higher hydrocarbons are converted to additional methane, CO₂ Removal Unit 113 for upgrading the methane to SNG quality, and SNG Drying Unit 114.

The hydrogen necessary for hydrotreating is produced in H₂ Recovery Unit 111 by absorption from a side stream of the purge gas from Methanol Synthesis Unit 110. The offgas of Unit 111 is recompressed to the SNG upgrading train pressure.

The H₂S contained in the Gasification Unit 105 offgas is converted in Sulphur Recovery Unit 106 to sulphur.

The gas liquors condensed in Units 101, 103, 104 and 105 are pumped to Gas Liquor Separation Unit 107, where dusty tar and oil are recovered. The gas liquor is processed further in Phenol Recovery Unit 108 and Ammonia Recovery Unit 109.

Sulphur and anhydrous ammonia are saleable products. Dusty tar, oil and phenols are not further upgraded, but are used as fuel.

3.4.2 Gasification Unit 101 (PFD ZO-GEM-6895 and 6896)

Purpose of the Unit is to convert coal into raw synthesis gas.

Technology Used

As defined in the scope of the study, the Lurgi Coal Pressure Gasification Process is used. This is a partial oxidation process by addition of oxygen and steam in the Lurgi countercurrent moving bed pressure gasifier.

Coal Handling Inside the Gasification Building

The well-sized coal is transported from Coal Handling Unit 128 to the top of the gasifiers by two-100 percent capacity inclined belt conveyors. During normal operation, each belt conveyor operates at 50 percent capacity and feeds two horizontal conveyors which run between two rows of gasifiers. Two spare horizontal conveyors are provided.

Coal is fed from the horizontal conveyors to the coal bunkers, 101-Y-01, by means of mobile automatic stations, each of which consists of a tripper and a reversible belt. Four stations have been provided with two stations normally operating.

All coal handling in the gasification building is controlled from a central control room.

Process Description

Gasification (PFD ZO-GEM-6895)

The well-sized coal is charged from the coal bunkers, 101-Y-01, into the gasifiers via automatically operated coal locks, 101-D-02. The bed of coal is gasified in countercurrent flow with the gasification agent. While travelling from the top to the bottom of the gasifier, the coal is dried, devolatilized and gasified.

The gasification agent is a mixture of oxygen and steam which is introduced into the bottom of the gasifier through a rotating grate. The ash is removed by this grate and discharged via the semi-automatic Ash Lock 101-D-03 and flows to Ash Handling Unit 129.

The steam addition moderates the gasifier temperature to below the ash liquefaction point. The partial combustion of the coal with the oxygen supplies the heat necessary for the gasification reactions.

The hot raw gas leaving the gasifier is cooled and scrubbed in Wash Cooler 101-E-01 with recycled gas liquor. By this operation the crude gas is saturated with steam and the dust and the heavy tars are removed. Before leaving the unit, further cooling takes place in Waste Heat Boiler 101-E-01 where 100 psig steam is produced.

The dusty gas liquor leaving the wash cooler and the waste heat boiler is sent to Gas Liquor Separation Unit 107.

Lock Gas Recompression Section of Unit 101 (PFD ZO-GEM-6896)

When the coal locks are depressured, raw gas is released. The portion which is released down to about 70 psig is fed to the last stage of Lock Gas Recompressor 101-C-41 after passing through Venturi Scrubber 101-D-41. The portion which is released down to atmospheric pressure is fed to the first stage of this recompressor via Lock Gas Holder 101-F-41.

The small quantity escaping during the charging operation of the coal locks, 101-D-02, is not recompressed due to the presence of air and is exhausted by Coal Lock Suction Fan 101-C-01 through Venturi Scrubber 101-F-05 to the atmosphere.

The lock gas recompressor also compresses the Gas Purification Unit 105 recompression gas and the main compressor syngas losses of Methanol Synthesis Unit 110 (Base Case I only).

3.4.3 Raw Gas Shift Unit 102 (PFD ZO--GEM-6897)

Purpose of the Unit is to increase the H_2/CO ratio of the raw synthesis gas to meet the requirements for methanol conversion. Only a sidestream of the raw gas has to be converted.

The exothermic shift reaction is as follows:

No additional steam is required for the reaction, since sufficient steam is contained in the raw gas.

In addition to the conversion of CO, most of the unsaturated hydrocarbons, higher phenols, HCN and organic sulfur are hydrogenated and the tar is hydrocracked into oil and naphtha.

Technology Used

For CO conversion in the raw gas generated in the Lurgi gasifier, Lurgi has developed a technology using commercial catalysts.

Process Description

The water content of the raw gas from Gasification Unit 101 is adjusted by cooling in the 50 psig Waste Steam Generator 102-E-04. In Separator 102-F-01, the condensed gas liquor is separated from the raw gas. The condensed gas liquor then flows to Gas Liquor Separation Unit 107 via Raw Gas Cooling Unit 103.

In Heat Exchangers 102-E-01, 02 and 03, the raw gas is heated to reaction temperature by heat exchange against the shifted gas. The raw gas then flows to Prereactor 102-D-01, which serves primarily as a guard bed trapping polymerized hydrocarbons and dust. The prereactor can be bypassed and regenerated when the pressure drop across the catalyst bed becomes excessive.

The partially shifted gas from the prereactor flows to Reactors 102-D-02 and 03 for further conversion. The shifted gas from 102-D-03 is cooled in Exchangers 102-E-01 and 02 against the raw feed gas and then flows to Shifted Gas Cooling Unit 104.

Provisions have been made to periodically regenerate the catalyst using steam and air.

3.4.4 Raw Gas Cooling Unit 103 (PFD ZO-GEM-6898)

Purpose of the Unit is to cool the raw synthesis gas coming from the gasifiers to a temperature as low as possible using cooling water, thereby reducing the downstream refrigeration requirements for gas purification.

Within the dictated temperature limitations, heat is recovered by the production of LP steam.

During cooling the raw gas, gas liquor, containing tar, oil, phenols and ammonia, is condensed.

Technology Used

The technology used has been developed by Lurgi for the special requirements of the Lurgi Coal Gasification Process.

Process Description

The raw gas is first cooled in the 50 psig Steam Generator 103-E-01. It is then cooled in the 20 psig Steam Generator 103-E-102. To further reduce the raw gas temperature, it is cooled by air in 103-E-03 and by cooling water in 103-E-04. To prevent the formation of ammonium carbonate, which could plug the tubes, gas liquor is injected into 103-E-04.

The gas liquor condensed in these steam generators and Trim Cooler 102-E-04 flows to Tarry Gas Liquor Tank 103-F-04. Part of the gas liquor is recycled to Wash Cooler 101-E-01 in Unit 101; the balance is sent to Gas Liquor Separation Unit 107.

3.4.5 Shifted Gas Cooling Unit 104 (PFD ZO-GEM-6899)

Purpose of the Unit is to cool the hot shifted gas to a temperature as low as possible using cooling water, thereby reducing the refrigeration requirements for gas purification. (See Paragraph 3.4.4).

Technology Used

The technology used has been developed by Lurgi for the special requirements of the Lurgi Coal Gasification Process.

Process Description

The shifted gas is first cooled in the 100 psig Steam Generator 104-E-01. The shifted gas is then cooled in the 20 psig Steam Generator 104-E-02. To further reduce the shifted gas temperature, it is cooled by air in 104-E-03 and by cooling water in 104-E-04. To prevent the formation of ammonium carbonates, which could plug the tubes, gas liquor is injected into 104-E-04.

After 104-E-04, the shifted gas is compressed by Booster Compressor 104-C-01 in order to overcome the pressure drop of Raw Gas Shift Unit 102. The compressed shifted gas is then cooled by 104-E-05 and combined with the cooled raw gas from Unit 103.

The gas liquor from the coolers and steam generators flows to Gas Liquor Separation Unit 107.

3.4.6 Gas Purification Unit 105 (PFD ZO-GEM-6900)

Purpose of the Unit is to remove the sulfur compounds and carbon dioxide from the raw gas to the desired levels required by the methanol conversion unit. Naphtha, hydrogen cyanide and water are also removed.

Technology Used

The selected Lurgi Rectisol Process purifies the raw gas by physical absorption using cold methanol as the absorption agent. The Rectisol process is currently in successful commercial operation. The application of this process is most economic at a high partial pressure of CO₂ in the feed gas.

Although a high selectivity of sulphur removal can be achieved to meet the requirements of the Claus sulphur recovery process, the non-selective route has been chosen using the Stretford sulphur recovery process. The Rectisol/Stretford combination is believed to be the most economic system for a low sulphur coal, based on previous studies for commercial-sized SNG plants.

The Naphtha and Methanol Recovery Section (NMR) of the Rectisol process has been developed to meet special requirements of the Lurgi Coal Gasification Process.

Process Description

The Refrigeration Unit 141 maintains the two temperature levels of -45°F and 32°F required for gas purification.

Absorber Section

The feed gas is cooled sequentially in 105-E-01 with synthesis gas, in 105-E-02 with + 32°F refrigerant, in 105-E-03 with synthesis gas and flash gases and in 105-E-04 with -45°F refrigerant. The feed gas, after cooling, enters the prewash section of Absorber 105-D-01 where it is washed with a small quantity of methanol to remove the naphtha, HCN, and water. The main section of the absorber consists of a low and a middle section, with an intermediate chilling loop, where the gas is scrubbed with flash regenerated methanol to remove the bulk of sulphur compounds and carbon dioxide. From the main section the gas flows to the fine wash section in the top of the absorber. In the fine wash section, the gas is scrubbed with hot regenerated methanol to remove the remaining sulphur compounds and carbon dioxide to the specified level for methanol synthesis. The gas then exits the unit by heat exchange with the feed gas, and the syngas flows to Methanol Synthesis Unit 110.

Regeneration Section

The rich methanol leaving the bottom of Absorber 105-D-01 is now regenerated in Flash Regenerator 105-D-02. The laden methanol is flashed in several steps down to a slight vacuum, where dissolved gases are released. Gas from the first flash stage is sent to Lock Gas Compressor 101-C-41 to recover the bulk of the co-absorbed CH₄, CO, and H₂. The other flash gases are sent to Sulphur Recovery Unit 106.

A portion of the bottoms of the flash regenerator becomes the semi-lean solution used in the middle section of the absorber, while the remainder is further regenerated in Hot Regenerator 105-D-03 by a steam heated reboiler. The fully regenerated methanol (lean methanol) is used in the top section of Absorber 105-D-01. The offgas leaving the hot regenerator is combined with the above described flash gases and sent to Sulphur Recovery Unit 106.

Naphtha and Methanol Recovery Section

The methanol from the prewash section of Absorber 105-D-01 is flashed in Prewash Flash Column 101-D-04 before it is sent to the naphtha and methanol recovery section, which is shown as a block on the PFD.

This block consists of: (1) Mixer-Settler Extractor for recovering of naphtha, (2) Azeotrope Tower for fine recovery of naphtha, and (3) Methanol Water Tower for recovering of the prewash methanol.

The extraction agent is the condensate recovered from the bottom of the methanol water tower and sent to Waste Water Treatment Unit 135.

The recovered naphtha, which contains all of the organic sulphur compounds and about 50% of the HCN contained in the raw gas is sent to Naphtha Hydrotreating Unit 151.

The remaining 50% of the HCN will leave the NMR section as a HCN offgas which is recycled by the Lock Gas Compressor 101-C-41 to Raw Gas Shift Unit 102 for conversion to ammonia.

3.4.7 Sulphur Recovery Unit 106 (PFD ZO-GEM-6901)

Purpose of the Unit is to recover the sulphur from the various acid gases of the plant complex. The sulphur is recovered as elemental sulphur with a purity of 99.5%.

Technology Used

The Stretford Process which was developed by the British Gas Corp. is used to generate salable, solid sulphur from low concentration H₂S gases. (These gases are not suited for the Claus process).

The Stretford Process is an absorption process using a sodium carbonate solution with additives.

The overall chemical reaction including absorption and regeneration is written as:

Process Description

The feed gas is contacted countercurrently in Absorption Column 106-D-01 by the circulating Stretford solution. The H₂S-free gas exits the top of the absorber and goes to Unit 122 to be incinerated. It contains less than 10 ppm of H₂S. None of the COS in the feed gas, however, is absorbed.

The rich Stretford solution enters Reaction Chamber 106-D-02 which provides the time necessary for the completion of the chemical reaction. Then, the rich Stretford solution flows by gravity to Oxidizer 106-D-03, where it is contacted with air. In 106-D-03, the Stretford solution is oxidized for reuse, and the sulphur particles float to the top of the liquid as a froth.

The regenerated Stretford solution overflows from the oxidizer into Solution Pump Tank 106-F-01, and, after cooling, it is recycled back to the absorber.

The sulphur froth from the oxidizer flows by gravity to Pump Tank 106-F-02, from which it is pumped to Sulphur Vacuum Filter 106-F-03. From there it flows by gravity to Sulphur Melter 106-E-02 and on to Melted Sulphur Accumulator 106-F-04.

3.4.8 Gas Liquor Separation Unit 107 (PFD ZO-GEM-6902)

Purpose of the Unit is to separate by gravity the dusty tar, oil and co-absorbed gases from the various gas liquor streams.

Technology Used

The technology used has been developed by Lurgi for the special requirements of the Lurgi Coal Gasification Process.

Process Description

Dusty Gas Liquor

The dusty gas liquor coming from the waste heat boiler in Unit 101 and from 103-F-04 in Unit 103 is cooled in 107-E-01 and mixed with the cooled gas liquor from Units 103 and 104. The mixture then flows into Tar Separator 107-F-01 via Expansion Vessel 107-F-02. In the tar separator, tar is settled out in the conical bottom section and pumped to Tank 107-F-10.

The gas liquor leaving 107-F-01 flows by gravity to Oil Separator 107-F-04. The oil separated in 107-F-04 flows by gravity to Oil Tank 107-F-08. The gas liquor flows by gravity to Surge Tank 107-F-07.

Oily Gas Liquor

The oily gas liquor coming from Heat Exchangers 104-E-01 and 02 of Shifted Gas Cooling Unit 104 is cooled and then mixed with the gas liquor coming from Gas Purification Unit 105. This combined stream flows by gravity, via Expansion Vessel 107-F-03, into Separator 107-F-05, where the separated oil flows by gravity to Oil Tank 107-F-08.

Gas Liquor

The gas liquor leaving Separator 107-F-05 flows by gravity to Final Separator 107-F-06, where it is combined with the gas liquor from Surge Tank 107-F-07.

The gas liquor separated in 107-F-06 flows by gravity to Final Surge Tank 107-F-09, and from there, it is pumped to Phenol Recovery Unit 108 for further treating. The oil recovered in 107-F-06 flows by gravity to Oil Tank 107-F-08.

Tar and Oil

The tar is burned in Boiler Unit 122. The oil is fuel for Superheaters 123 and 124. The various expansion gases are scrubbed in Ammonia Scrubber 107-D-01 for ammonia recovery and then sent to Sulphur Recovery Unit 106.

3.4.9 Phenol Recovery Unit 108 (PFD ZO-GEM-6903)

Purpose of the Unit is to recover phenolic components from the gas liquor.

Technology Used

For the removal of phenols, the Lurgi Phenosolvan Process is provided. It is a solvent extraction process using di-isopropyl ether as the solvent. The Phenosolvan Process has been commercially used in syngas and town gas plants based on Lurgi Pressure Coal Gasification and is under consideration for the proposed U.S. SNG plants.

Process Description

The gas liquor coming from Gas Liquor Separation Unit 107 flows to Sand Filters 108-F-02, where any remaining solids are removed. The filters are backwashed with cleaned gas liquor and the mud liquor from the filters is returned to Unit 107. From the sand filters, the cleaned gas liquor flows to Saturation Column 108-D-01, where it is saturated with CO₂ in order to increase the efficiency of extraction.

The saturated gas liquor flows by gravity to the multi-stage, mixer settler type extractor, 108-D-02, to be divided into a phenolic-solvent phase and a dephenolized-gas-liquor phase.

The dephenolized gas liquor flows to Raffinate Settler 108-F-06 for the purpose of recovering the remaining solvent. From there the gas liquor is pumped to Ammonia Recovery Unit 109.

The phenolic-solvent phase from the extractor flows, via Extract Settler 108-F-07, to Extract Receiver 108-F-08. It then is pumped, via Partial-Feed Evaporator 108-E-02, into Solvent Regenerator 108-D-04. In the overhead of this column, the bulk of the solvent is recovered and pumped back to Extractor 108-D-02. The bottom product is a phenol-solvent mixture which is pumped into Phenol Tower 108-D-05, where the remaining solvent is stripped out using live steam.

The overhead solvent-water mixture of 108-D-05 is returned to Extract Settler 108-F-07. The bottom product of phenols is pumped to Main Boiler Unit 122 to be burned.

A small portion of the phenols is recycled, partly for recovering solvent from the offgases in Vent Scrubber 108-D-06 and partly for recovering solvent from Unit 109 offgases in Phenol Solvent Scrubber 108-D-03. In the upper part of the phenol solvent scrubber, these offgases are scrubbed by waste water for phenol recovery, before being sent to Sulphur Recovery Unit 106.

3.4.10 Ammonia Recovery Unit 109 (PFD ZO-GEM-6904)

Purpose of the Unit is to recover the ammonia from the gas liquor before final treatment for use as cooling water make-up.

Technology Used

The technology used has been jointly developed by Chemie Linz and Lurgi, mainly for the purpose of removing ammonia from the Lurgi gasifier gas liquor. The product is a salable ammonia of either agriculture or chemical grade.

Ammonia recovery in the Chemie Linz/Lurgi Process is accomplished by the use of stripping and scrubbing techniques without the addition of chemicals or solvents.

Process Description

After heating, the gas liquor enters Deacidifier 109-D-01, where the remaining gases and phenosolvan solvent are stripped out by reboiler steam. The gases leaving the stripper are scrubbed with cooled waste water under elevated pressure to remove ammonia. The stripped gases are sent back to Phenol Recovery Unit 108. The gas liquor containing the ammonia leaves the bottom of the deacidifier and is pumped after cooling to Total Stripper 109-D-02, where all the ammonia and the remaining gases are stripped out by reboiler steam under atmospheric pressure.

The bottom product of the total stripper is the clean gas liquor containing less than 100 ppm free ammonia. It is sent after cooling to Waste Water Treatment Unit 135 for final treating.

The overhead vapors from the total stripper are partially condensed in the bottom of Acid Gas Scrubber 109-D-03 in order to increase the ammonia concentration. The condensate is refluxed back to the total stripper.

In Acid Gas Scrubber 109-D-03, the vapors are scrubbed by a recycled aqueous ammonia solution from the process. The heats of reaction and condensation are removed by cooled pump-arounds.

The bottom stream of the acid gas scrubber, which contains free ammonia and ammonium salts, is pumped by 109-G-05 to Ammonia Stripper 109-D-04. The bottoms of the ammonia stripper, containing a highly concentrated stoichiometric ammonium salt solution, is recycled back to 109-D-01. The stripped ammonia from the ammonia stripper overhead flows to the acid gas scrubber for recovery.

The ammonia vapor flowing overhead from Acid Gas Scrubber 109-D-03 is condensed and dried by compressing and cooling in the Ammonia Liquefying and Drying (ALD) Section before being sent to storage.

A small liquid stream containing water and organic substances is removed in the ALD Section and sent to the main waste water stream.

3.4.11 Methanol Synthesis Unit 110 (PFD ZO-GEM-6905)

Purpose of the Unit is to produce methanol from the purified synthesis gas.

Technology Used

For the production of methanol, the Lurgi Low Pressure Methanol Synthesis Process has been selected. This process offers maximum safe operation due to its highly developed reactor design and has wide commercial usage.

Because of the high methane content, syngas conversion is:

H ₂	=	82.0 mol %
CO	=	93.1 mol %

Process Description

The synthesis gas is compressed, Syngas Compressor 110-C-01, to about 1100 psig and mixed with the recycle gas stream. The Recycle Gas Compressor 110-C-02 compresses the mixture to overcome the pressure losses of the recycle loop. This loop consists of Heat Exchanger 110-E-03 (heating syngas/recycle gas against methanol/recycle gas), Reactor 110-D-01, Condensers 110-E-04 and 05, and KO Drum 110-F-01, in which the unconverted gas is separated for recycling.

In Reactor 110, the heat of reaction is transferred to boiler feed water flowing outside the catalyst tubes, thereby generating medium pressure steam. The reactor temperature is controlled by controlling the steam pressure.

The high pressure methanol liquid is expanded into Expansion Vessel 110-F-03 for degassing; the methanol is sent to Methanol Conversion Unit 150.

The expansion gas of 110-F-03 is recompressed in H₂ Recovery Unit 111 and sent to Methanation Unit 112. To eliminate the build-up of methane and inerts in the recycle gas, a portion of the recycle gas is purged continuously and fed to Methanation Unit 112.

3.4.12 H₂ Recovery Unit 111 (PFD ZO-GEM-6908)

Purpose of the Unit is to recover hydrogen for hydrotreating the gasifier naphtha. The hydrogen source is a slip stream from the methanol unit purge gas.

Technology Used

Because of the relatively high CO₂ content of the feedgas, the Pressure Swing Process of Union Carbide Corp. is used.

Process Description

The main equipment of the unit are four parallel pressure-swing absorbers, 111-D-01; one is in operation and three are in regeneration. During operation, essentially all of the non-hydrogen components and a portion of the hydrogen are absorbed. The remaining hydrogen leaves the absorber at a purity of over 98%.

Regeneration is by pressure reduction in 3-stages with the final pressure being nearly atmospheric. The regeneration cycle is controlled automatically by motor valves. An absorption/regeneration cycle requires several minutes.

The desorbed gases, called pressure-swing offgases, are recompressed by Compressor 111-C-01 back into the methanol unit purge gas stream before being charged to Methanation Unit 112. Gas Holder 111-F-01 on the suction side of the compressor, is necessary for buffering the non-static, pressure-swing off gas flow from the regenerating absorbers. Compressor 111-C-01 also recompresses the expansion gas from Unit 110.

3.4.13 Methanation Unit 112 (PFD ZO-GEM-6906)

Purpose of the Unit is the production of SNG meeting the specifications for compatibility with existing pipeline gases. In addition, the CO content must be lowered to less than 0.1%.

The principal reactions in methanation are described by the following equations:

Other minor reactions include the hydrogenation of ethylene to ethane and the cracking of ethane to methane.

Technology Used

The Lurgi Methanation Process is used. This process is the first process developed for the methanation of high CO concentration gases, and has been included in the proposed commercial U.S. SNG plants.

Process Description

The feed gas is made up of the purge and expansion gases from Methanol Synthesis Unit 110.

The methanation unit consists of two down-flow, fixed-bed reactors in series. The feed gas is heated in 112-E-06 and 03 against the second reactor, 112-D-02, product. Steam is injected in the feed gas and the gas is also diluted with a portion of the methanated gas from the first reactor, 112-D-01. The dilution gas is compressed by Recycle Compressor 112-C-01, after having been cooled in 112-E-01 and 02. MP steam is generated in 112-E-01.

The net reactor effluent from the first reactor flows to the second, or clean-up, reactor in which final methanation takes place. The clean-up reactor effluent is heat exchanged against the feed and cooled in Exchangers 112-E-04 and 05.

Process water condensed from the reactor effluent stream is collected and routed to BFW Preparation Unit 131.

The methanated gas leaves the unit for further treating in CO₂ Removal Unit 113.

3.4.14 CO₂ Removal Unit 113 (PFD ZO-GEM-6907)

Purpose of the Unit is to lower the CO₂ content in the SNG to about 0.5%, thereby improving its quality.

Technology Used

The Amine Guard Process of Union Carbide Corp. uses a monoethanol amine solution as the absorbant. This process offers better economics than a conventionally designed MEA wash system and provides complete protection against corrosion. A MEA process was chosen because the specified low final CO₂ content does not allow the use of a hot potassium carbonate process and the partial pressure of CO₂ in the feed gas is too low for economical use of a physical absorption process, such as Rectisol.

Process Description

The Methanation Unit 112 product gas and the offgases of Fractionation Unit 152 are charged to Absorber 113-D-01 in which the CO₂ is washed out by counter-current flow with lean MEA solution. In the top of the absorber, the gas is scrubbed with a small flow of BFW to recover trace amounts of MEA. The purified gas then flows to SNG Drying Unit 114.

The absorber bottoms product (rich MEA solution) is expanded after heat exchange in 113-E-01 into MEA Regenerator 113-D-02, where the CO₂ and other co-absorbed compounds are stripped from the MEA solution. The steam in the regenerator overhead is condensed in Condensers 113-E-04 and 05, and the remaining offgases are sent to Boiler Unit 122 for incineration. The steam condensate is refluxed to the top of the regenerator. The lean solution from the bottom of the regenerator is pumped back to the Absorber 113-D-01, after being cooled in 113-E-01, 02 and 03.

The offgases of Fractionation Unit 152 are compressed to the SNG pressure by Light Gas Compressor 113-C-01. After compression, the gas is cooled and the heavier hydrocarbons, which condense, are returned to Unit 152.

3.4.15 SNG Drying Unit 114 (PFD ZO-GEM-6907)

Purpose of the Unit is to dry the wet SNG to a water content of 4 lb/MMSCF before its introduction into the SNG pipeline system.

Technology Used

The TEG Wash Process, a commercial process for drying natural gas, is used.

Process Description

The drying agent is a mixture of triethylene glycol (TEG) and other ethylene glycols.

The SNG is contacted countercurrently in Water Absorber 114-D-01 with lean TEG. The water-rich TEG flows from the absorber to TEG Regenerator 114-D-02, which is heated by fuel gas. The steam stripped from the TEG is discharged to the atmosphere. The small reflux stream required is condensed within the vent tube by air cooling.

The regenerated TEG is stored in 114-F-01, where it is cooled by the rich TEG, before being pumped into the top of Water Absorber 114-D-01.

3.4.16 Methanol Conversion Unit 150 (PFD ZO-GEM-6915)

Purpose of the Unit is to dehydrate crude methanol to LPG and gasoline hydrocarbons which can be marketed directly without further processing.

Technology Used

Mobil has demonstrated a simple catalytic process to dehydrate methanol to the stoichiometric yield of hydrocarbons (44%) and water (56%).

This development has been reported under DOE Contract No. E(49-18)-1773, and in references cited. The hydrocarbons are predominately in the gasoline boiling range (C_4 to C_{10}), and the gasoline is chemically conventional consisting of highly branched paraffins (51%), highly branched olefins (13%), naphthenes (8%) and aromatics (28%). Its unleaded research octane number (RON) ranges from 90 to 100. Essentially, no hydrocarbons larger than C_{10} and no oxygenates are produced.

Small amounts of CO, CO₂ and coke and trace amounts of formic acid and acetone in the water product are formed as by-products. After fourteen days on stream, enough coke accumulates on the catalyst to require regeneration. The catalyst is regenerated by coke combustion in a controlled manner to limit the maximum temperature to 900°F. Since the catalyst converts other oxygenates and is tolerant to water, crude methanol without any purification can be charged directly, thereby simplifying the methanol synthesis.

From laboratory investigations, the reaction path from methanol to hydrocarbons appears to be reasonably well represented by:

The reaction is highly exothermic with the heat of reaction being about 740 Btu/lb of methanol. Therefore, the principal problem in designing a reactor system is heat removal. In the fixed bed reactor system, this problem is minimized by dividing the above overall reaction (and reactor system) into two steps. In the first, methanol is partially dehydrogenated to form an equilibrium mixture of methanol, dimethyl ether and water. About 20% of the heat of reaction is released in this first step without any recycle for temperature control. In the second step, where hydrocarbon conversion takes place, a large recycle gas stream (from 6/1 to 9/1 ratio) is used to limit the temperature rise across the bed to about 125°F.

While other systems, such as quenched reactor, tubular heat exchange reactor, or fluidized bed reactor could be designed, the fixed bed, gas recycle system has been well demonstrated and is readily scaleable to higher throughputs.

The commercial process design conditions, yields and catalyst properties used in this study are shown in Tables 3.4.1, 3.4.2 and 3.4.3, respectively. These data were developed from DOE Contract No. E(49-18)-1773 and represent the best estimate at this time of a commercial fixed bed operation.

TABLE 3.4.1

PROCESS DESIGN CONDITIONS FOR FIXED BED
METHANOL CONVERSION UNIT 150

I. Nominal Operating ConditionsFeed Composition (See Material Balance Stream 37.)

Methanol	18,360.3 lb. mol/hr.
Water	1,382.4 lb. mol/hr.
Others	42.5 lb. mol/hr.
	<u>19,785.2 lb. mol/hr. (52,585 B/SD)</u>

Base Yields (See following Table 3.4.2.)*Reactor ConditionsFirst Reactors (DME)

Pressure, Inlet	315 psig
Temperature, Inlet	566°F
Temperature, Outlet	768°F
Space Velocity	6 lb. fresh feed/hr/lb. catalyst
Catalyst	Mobil first stage catalyst
Estimated Ultimate Catalyst Life	2 years

Second Reactors (M-Gasoline)

Pressure, Inlet	300 psig
Temperature, Inlet	625°F
Temperature, Outlet	752°F
Space Velocity	1.84 lb. fresh feed/hr/lb. catalyst
Molar Recycle Ratio	7.5/1 (based on fresh feed)
Catalyst	Mobil second stage catalyst
Estimated On-Stream Time Before Regeneration	14 days
Estimated Ultimate Catalyst Life	1 year

II. Regeneration Conditions

Aged Catalyst Coke Level (Second Reactors) 30 wt. % on catalyst (Controlled oxygen content and controlled maximum temperature procedure similar to conventional petroleum naphtha pretreater/reformer regeneration is employed.)

TABLE 3.4.1 (continued)

III. Metallurgy

The basic metallurgy for the Methanol Conversion Unit 150 is carbon steel. Reactors have a 304SS internal liner over 3" of refractory to minimize the effects of any possible temperature excursions. The high pressure separator, 150-F-01 in process flow diagram ZO-GEM-6915, has a PVC lining to protect against the acid water (pH=3.5). The tubes of Exchangers 150-E-01, 06 and 07 are 316SS to protect from acid water corrosion. In addition, the shell of Exchanger 150-E-01 is carbon steel with 316SS lining. The acid water lines are PVC lined.

- * Yields are the average over 14 days of operation before catalyst regeneration. Although yield shifts are significant during the 14 days (see Reference 2), the 5-reactor system, Unit 150 design basis, minimizes the yield shift effects.

TABLE 3.4.2

METHANOL CONVERSION UNIT 150
 BASE FEED AND PRODUCT COMPOSITION
 (Basis: 100 Lb-Mole Methanol in Feed)

Component Names	Feed	Product
Coke (as CH _{0.8})	.000	.128
Acetone	.000	.129
Formic Acid	.000	.163
Methanol	100.000	.000
Dimethylether	.000	.000
Water	7.529	106.951
Carbon Monoxide	.000	.017
Carbon Dioxide	.000	.053
Hydrogen	.000	.040
Methane	.000	.746
Ethane	.000	.193
Ethene	.000	.018
Propane	.000	1.474
Propene	.000	.064
N-Butane	.000	.665
I-Butane	.000	2.155
Butenes	.000	.270
N-Pentane	.000	.268
I-Pentanes	.000	2.353
Pentenenes	.000	.435
Cyclopentane	.000	.047
Methylcyclopentane	.000	.214
N-Hexane	.000	.113
I-Hexanes	.000	1.991
Hexenes	.000	.297
Methylcyclohexane	.000	.062
N-Heptane	.000	.026
I-Heptanes	.000	.786
Heptenes	.000	.292
1,3-Dicyclopentane, cis	.000	.233
I-Octanes	.000	.228
Octenes	.000	.301
N-Propylcyclopentane	.000	.299
N-Nonane	.000	.015
I-Nonanes	.000	.084
Nonenes	.000	.116
N-Butylcyclopentane	.000	.071
I-Decanes	.000	.024
Decenes	.000	.045
Benzene	.000	.036
Toluene	.000	.280
Ethylbenzene	.000	.070
m- + p-Xylenes	.000	.876
o-Xylene	.000	.240
1,2,4-Trimethylbenzene	.000	.818
1,3,5-Trimethylbenzene	.000	.034

TABLE 3.4.2 (continued)

Component Names	Feed	Product
p-Ethyltoluene	.000	.292
I-Propylbenzene	.000	.014
1,2,4,5-Tetramethylbenzene	.000	.436
1,2,3,5-Tetramethylbenzene	.000	.063
1,2,3,4-Tetramethylbenzene	.000	.023
p-Diethylbenzene	.000	.198
Penta-Methylbenzene	.000	.068
2-Methylnaphthalene	.000	.017
Total Lb-Mole	107.529	124.832
Weight (lbs)	3,339.846	3,339.846

TABLE 3.4.3
METHANOL CONVERSION UNIT 150
CATALYST PROPERTIES

Type	-----Mobil Catalysts-----	
	<u>First Stage</u>	<u>Second Stage</u>
	Extrudate	Extrudate
Geometry	1/16" x 3/16"	1/25" x 3/16"
Bulk Density	35 lb/cu ft	35 lb/cu ft
Crush Strength	50 lb/in	40-50 lb/in
Cost	*	*

- * Since Mobil has no specific commercial experience in making the second stage catalyst in the amounts required for a major syn fuels industry, an accurate cost cannot be given at this time. We are sure, however, that its cost will contribute only a minor portion to the total gasoline cost and, therefore, its addition at a later time will not alter any current economic evaluation study conclusions.

Process Description

Depressured crude methanol, pumped (150-G-01) to reaction pressure and vaporized in Heat Exchangers 150-E-01 and 02, flows to DME Reactor 150-D-01, where it is catalytically converted to an equilibrium mixture of methanol, dimethylether and water. Inlet operating conditions are 315 psig and 566°F.

The equilibrium mixture then flows to the M-Gasoline reactors (150-D-02), of which five are operating and one is undergoing regeneration. A gas recycle stream from Separator 150-F-01, after compression (150-C-01) and heat exchanged (150-E-04 and 05) against the reactor effluent, is fed along with the equilibrium mixture into the M-Gasoline reactors. The recycle stream controls the temperature rise. Its temperature is maintained to give a reactor inlet temperature of 625°F.

Cooling of Reactor 150-D-02 effluent occurs as follows: (1) Steam Generator 150-E-03, (2) the recycle stream and methanol feed heat exchangers, (3) BFW Heat Exchanger 150-E-06 and (4) Air Trim Cooler 150-E-07. Separator 150-F-01 operating conditions are 248 psig and 155°F. The purge gas flow is limited to that only required for control. The liquid hydrocarbon product is pumped (150-G-02) to Unit 152. A small amount, five gpm, of hydrocarbon product is recycled from the separator to the inlet of BFW Exchanger 150-E-06 for durenene crystallization control.

Regeneration of the M-Gasoline reactors takes place about once every 14 days. After a reactor is taken off line, it is depressured and purged with nitrogen until there is less than about 1 to 5 vol. % of hydrocarbons present. Then the reactor is pressured to about 150 psig with nitrogen flow through Recycle Compressor 150-C-02. The temperature is maintained at 650°F by use of Furnace 150-B-01. Air is introduced slowly until oxygen content reaches about 1 vol. %. The maximum catalyst bed temperature is maintained at less than 850°F. When the oxygen content in the recycle gas becomes higher than 0.1 vol. %, the inlet temperature is slowly raised while maintaining a maximum temperature of 900°F in the bed. The hot spent regeneration gas is cooled by heat exchange with the fresh regeneration gas (150-E-08) and Fan Cooler 150-E-09. Condensed water is removed via Drum 150-F-02.

3.4.17 Naphtha Hydrotreating Unit 151 (PFD ZO-GEM-6908)

Purpose of the Unit is to obtain a clean, odor free, stable material from the raw gasifier naphtha for gasoline pool blending. Reaction conditions chosen will substantially saturate the raw naphtha olefins with little aromatic-ring hydrogenation, thereby reducing the octane rating as little as possible.

Technology Used

Hydrotreating is a versatile process in which the quality of a wide variety of petroleum materials can be improved with the proper choice of catalyst and operating conditions. It is used commercially to remove impurities such as sulfur, nitrogen and oxygen, to saturate olefinic double-bonds and to improve the odor, color, stability and burning characteristics. Generally, catalysts and conditions employed result in as little as possible carbon-bond cracking. Pressures range from 200 to 1,000 psig, and temperatures from 400 to 750°F. Typical liquid hourly space velocities are from 1 to 10 volume of oil per hour per volume of catalyst. Hydrogen-rich gas recycle may or may not be used depending upon the severity of operation. Since hydrogenation reactions are exothermic, multi-bed reactors with a hydrogen recycle quench are used to control reactor temperatures when the hydrogen consumption is large. A large number of catalysts have been developed. Various cobalt-molybdena catalysts with alumina supports; however, account for the bulk of the applications. This catalyst is oxygen regenerable, and quite rugged in nature. Commercial processes are licensed by BP Trading, IFP, Exxon, Gulf, Shell, Union Oil and Standard Oil (Indiana).

Process Description

The raw naphtha is pumped (151-G-01) to reactor pressure, heat exchanged with the reactor effluent in Exchanger 151-E-01, heated to reactor temperature in Fired Heater 151-B-01, and charged to Reactor 151-D-01. A portion of the hydrogen feed is mixed and heated with the naphtha; the remainder is used as a cold quench for temperature control in the multi-bed reactor. The reactor effluent is further cooled (151-E-02 and 03) before flushing in Separator 151-F-01. A small BFW flushing stream is added before 151-E-01.

The hydrogen-rich vapor passes first through an amine tower for hydrogen sulfide control, and then split into recycle and purge gas streams. The purge gas is incinerated in Boiler Unit 122. After compressing (151-C-01), the recycle and hydrogen make-up are combined before separating into the hydrogen feed and quench streams.

The separator liquid product is charged to Stabilizer 151-D-05. The light gas/hydrogen sulfide overhead products are sent to Sulfur Recovery Unit 106. Condensed Accumulator 151-F-03 liquids are refluxed. Cooled (151-E-06 and 07) stabilized naphtha is sent to offsite Blending Unit 154. A small sour water stream is also obtained from Separator 151-F-01.

Regeneration systems are provided for Reactor 151-D-01 and the rich amine solution from Absorber 151-D-02.

3.4.18 Fractionation Unit 152 (PFD ZO-GEM-6916)

Purpose of the Unit is to obtain a low C₂ concentration alkylate unit charge and a stabilized gasoline for pool blending.

Process Description

The high pressure condensate (stream 58) from the SNG train is flashed in Drum 150-F-01. The depressured liquid and the hydrocarbon product from Unit 150 are fed to Deethanizer 152-D-01. The ethane-rich tower overhead and the flash drum vapors are sent to Unit 113 in the SNG train. The deethanizer bottoms after being cooled in Deethanizer Side Reboiler 152-E-02 is fed to Stabilizer 152-D-02. Products are cooled alkylation feed (overhead) and stabilized gasoline (bottoms). Both towers employ steam reboilers (152-E-01 and 05).

3.4.19 Alkylation Unit 153 (PFD ZO-GEM-6916)

Purpose of the Unit is to increase the gasoline yield by the reaction between iso-butane and butene/propene and to yield commercial grade propane and butane liquified petroleum gases.

Technology Used

Alkylation, in petroleum refining, normally means the reaction of butene and/or propene with iso-butane to produce a higher molecular weight iso-paraffin. For example:

Alkylation can occur without catalysts, but commercially, acid catalysts are used with sulfuric acid and hydrogen fluoride being the two most important. Unlike the simple reaction above, a mixture of isomers is obtained. In addition, polymerization occurs, and "sludge" is yielded requiring catalyst regeneration. Alkylation, highly exothermic, liberates 600 to 700 Btu/lb of iso-butane. Commercial processes are licensed by Phillips Petroleum, UOP Process and Stretford Engineering.

Process Description

The iso-butane/olefin fresh feed is first mixed with the cooled HF acid/iso-butane recycle streams (153-E-16), and then, this mixture is charged to the combined Reactor/Settler 153-D-01. The reactor effluent - inert hydrocarbons, excess iso-butane and alkylate - after heating (153-E-01 and 02) is fed to Deisobutanizer 153-D-02. Products are a C₃/i-C₄ overhead, mixed-butanesside-cut and alkylate bottoms. The overhead mixture, after condensing in Condensers 153-E-08 and 09, is pumped (153-G-01) into Depropanizer 153-D-03 for separation into the C₃ LPG product and excess iso-butane recycle.

The hydrogen fluoride is recycled from Reactor/Settler 153-D-01. A continuous slip stream from the acid recycle is purified in a conventional HF rerun unit (not shown).

Not shown on the simplified process flow diagram are a HF stripper in the C₃ LPG stream and caustic washes in the product streams to remove all traces of acid.

3.4.20 Unit Capacity Summary

The capacities of the principal process units have been summarized below in conventional flow rates:

<u>Unit No.</u>	<u>Name</u>	<u>---Capacity, Unit/SD---</u>	
		<u>Input</u>	<u>Output</u>
101	Gasification	22.8 MT coal	1,755 MMSCF
102	Raw Gas Shift	331 MMSCF	-
105	Gas Purification	1,062 MMSCF	742 MMSCF
110	Methanol Synthesis	742 MMSCF	53.3 M Bbl
150	Methanol Conversion	52.6 M Bbl	25.7 M Bbl
151	Naphtha Hydrotreating	1.3 M Bbl	-
152	Fractionation	25.7 M Bbl	-
153	Alkylation	5.3 M Bbl (750 Bbl/SD alkylate)	-
	SNG Train	213 MMSCF	148 MMSCF

1) The steam/BFW circuit of the complex is shown simplified as it is described in detail within the separate steam/BFW balance Z0-GEM-6910.

GEM		Lurgi	
00-0865		Lurgi Mineraltechnik GmbH	
BLOCK FLOW DIAGRAM		Z0-GEM-6894	
COAL TO GASOLINE CONV.		WRC/LRDA STUD	
VIA METHANOL		EF-76-C-01-247	
BASE CASE I

EQUIPMENT LIST			
ITEM NO.	NAME	QPM	APPROX. -SPACE (D. x H) (Ft. x Ft.)
101-C-41	LOCK GAS COMPRESSOR	1 x 0	
101-B-41	LOCK GAS VENTURI SCRUBBER	1 x 0 4'-0"	30'
101-F-41	LOCK GAS BOLLER	1 x 0	
101-G-41	LOCK GAS SCRUBBER P/A PUMP	1 x 1	

◇	STREAM NUMBER	
○	PRESSURE	PSI
□	TEMPERATURE	°F
▭	FLOW RATE DRY GAS	MMSCFD
▭	FLOW RATE STEAM AND LIQUIDS	MMBtu/hr

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE

THIS PFD IS IDENTICAL FOR BASE CASE I AND II
EXCEPT AS NOTED

GEM 00-0865		LURGI Lurgi Motorchemisches GmbH Frankfurt (Main)	
PROCESS FLOW DIAGRAM RECOMPRESSION SECTION OF THE GASIFICATION UNIT		Z0-GEM-6896 METHANOL STUDY EF-70-C-01-247	
UNIT 101/201		BASE CASE I/II

EQUIPMENT LIST		
ITEM NO.	NAME	OPER. APPROX. + SPARE I.D. LENGTH T-F
102-B-01	SHIFT START-UP HEATER	1 x 0
102-D-01	SHIFT PREHEATER	2 x 0 11'-0" SPHERE
102-D-02	FIRST SHIFT REACTOR	2 x 0 13'-0" 34"
102-D-03	SECOND SHIFT REACTOR	2 x 0 13'-0" 34"
102-E-01	FIRST SHIFT H.E.	2 x 0
102-E-02	SECOND SHIFT H.E.	2 x 0
102-E-03	THIRD SHIFT H.E.	2 x 0
102-E-04	50 PSIG STEAM GENERATOR	2 x 0
102-F-01	PLTD GAS LIQ. SEPARATOR	2 x 0
102-F-02	50 PSIG STEAM HEAT	2 x 0

◇	STREAM NUMBER	
○	PHASE	PSIG
□	TEMPERATURE	°F
▱	FLOW RATE DRY GAS	MMBtu/hr
▭	FLOW RATE STEAM AND LIQUIDS	MMBtu/hr

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE

THIS PFD IS IDENTICAL FOR BASE CASE I AND II
EXCEPT AS NOTED

GEM 00-0865		LURGI Lurgi Mineralogische GmbH Frankfurt, Germany	
PROCESS FLOW DIAGRAM RAW GAS SHIFT		Z0-GEM-6897	
UNIT 102/202		BASE CASE I/II

ITEM NO.	NAME	OPER. - -SPARE
103-E-01	50 PSIG STEAM GENERATOR	3 + 0
103-E-02	20 PSIG STEAM GENERATOR	3 + 0
103-E-03	COOLER	3 + 0
103-E-04	TRIM COOLER	3 + 0
103-F-01	50 DRUM	3 + 0
103-F-02	50 PSIG STEAM DRUM	3 + 0
103-F-03	20 PSIG STEAM DRUM	3 + 0
103-F-04	TARRY GAS LIQUOR TANK	1 + 0
103-G-01	TARRY GAS LIQUOR PUMP	1 + 1

◇	STREAM NUMBER	
○	PRESSURE	PSIG
□	TEMPERATURE	°F
▱	FLOW RATE DRY GAS	Standard
▭	FLOW RATE STEAM	Standard
▮	AND LIQUIDS	Standard

NOTES:
THIS PFD IS IDENTICAL FOR
BASE CASE I AND II

STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE

GEM 00-0865		Lurgi Long-Range-Process-Design	
PROCESS FLOW DIAGRAM RAW GAS COOLING		Z0-GEM-6898	
UNIT 103/203		BASE CASE 1/2

EQUIPMENT LIST		
ITEM NO.	NAME	OPER. + SPARE
104-C-01	BOOSTER COMPRESSOR	1 + 0
104-E-01	100 PSIG STEAM GENERATOR	1 + 0
104-E-02	20 PSIG STEAM GENERATOR	1 + 0
104-E-03	COOLER	1 + 0
104-E-04	TRIP COOLER	1 + 0
104-E-05	FINAL COOLER	1 + 0
104-F-01	FIRST E.O. DRUM	1 + 0
104-F-02	SECOND E.O. DRUM	1 + 0
104-F-03	100 PSIG STEAM DRUM	1 + 0
104-F-04	20 PSIG STEAM DRUM	1 + 0

◇	STEAM NUMBER	
○	PSI	PSI
□	TEMPERATURE	°F
▭	FLOW RATE DRY GAS	Normal / hr
▭	FLOW RATE STEAM AND LIQUIDS	Wt / hr

NOTES:
THIS PFD IS IDENTICAL FOR
BASE CASE I AND II

STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE

GEM		LURGI	
00-0865		Lurgi Mineraltechnik GmbH	
PROCESS FLOW DIAGRAM		Z0-GEM-6899	
SHIFTED GAS COOLING		REVISION STUDY	
A0		DATE: 01-24-47	
UNIT 104/204		BASE CASE I/II

ITEM NO.	NAME	EQUIPMENT LIST	OPER. APPROX. *SPARE I.D. LENGTH Y.
106-C-01	AIR BLOWER		2 + 0
106-D-01	ABSORBER		8 + 0 17'-0" / 810"
106-D-02	REACTION CHAMBER		2 + 0 20'-0" / 32"
106-D-03	OXIDIZER		2 + 0 56'-0" / 28"
106-E-01	SOLVENT HEATER		2 + 0
106-E-02	SULPHUR MELTER		2 + 0
106-F-01	SOLUTION PUMPING TANK		2 + 0
106-F-02	SLURRY TANK		2 + 0
106-F-03	SULPHUR VACUUM FILTER		2 + 0
106-F-04	WETTED SULPHUR ACCUMULATOR		2 + 0
106-G-01	LEAN SOLUTION PUMP		2 + 1
106-G-02	SLURRY PUMP		2 + 1
106-G-03	RECYCLE PUMP		2 + 1
106-G-04	WETTED SULPHUR PUMP		1 + 1
106-G-05	RICH SOLUTION INTERMEDIATE PUMP		4 + 2

◇	STREAM NUMBER	
○	PRESSURE	PSIG
□	TEMPERATURE	°F
▱	FLOW RATE GPM GAS	SCFM
▭	FLOW RATE STEAM AND LIQUIDS	MMBtu/hr

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE.
THIS PFD IS IDENTICAL FOR
BASE CASE I AND II

Project	28777 VI	GEM	00-0865	IURGI
Client				Lurgi Minnaphos GmbH
Design				From 1st Meeting
Drawn by				ZO-GEM-6901
Checked by				MPDC EPLA STUDY
Approved by				EF-75-C-01-2447
Scale				
Unit	106 / 206	BASE CASE I/II

ITEM NO.	NAME	EQUIPMENT LIST	OPER. APPROX. -SPARE I.D. LENGTH TYPE
108-D-01	SATURATOR		1 x 0 10'-0" 40'
108-D-02	EXTRACTOR		2 x 0 MIXER SETTLER
108-D-03	SCRUBBER		1 x 0 3'-0" 81'
108-D-04	REGENERATOR		1 x 0 15'-0" 63'
108-D-05	TOWER		1 x 0 5'-0" 68'
108-D-06	VENT SCRUBBER		1 x 0 3'-0" 50'
108-E-01	PLATE LIQUOR HEATER		1 x 0
108-E-02	REGENERATOR FEED EVAPORATOR		1 x 0
108-E-03	REGENERATOR REBOILER		1 x 0
108-E-04	REGENERATOR OVER CONDENSER		1 x 0
108-E-05	REGENERATOR OVER TRIM CONDENSER		1 x 0
108-E-06	TOWER REBOILER		1 x 0
108-E-07	RECTIFY CONDENSER		1 x 0
108-E-08	TOWER OVER CONDENSER		1 x 0
108-F-01	FEED STORAGE TANK		1 x 0
108-F-02	SAND FILTER		8 x 1
108-F-03	SAND FILTER PLATE LIQUOR TANK		1 x 0
108-F-04	SAND FILTER AND LIQUOR TANK		1 x 0
108-F-05	TAR/OIL SLOP RECEIVER		1 x 0
108-F-06	RAFFINATE FINAL SETTLER		2 x 0
108-F-07	EXTRACT SETTLER		2 x 0
108-F-08	EXTRACT RECEIVER		2 x 0
108-F-09	ACID GAS E.O. DRUM		1 x 0
108-F-10	REGENERATOR OVER SETTLER		1 x 0
108-F-11	REGENERATOR OVER ACCUMULATOR		1 x 0
108-F-12	PRODUCT SURGE DRUM		1 x 0
108-G-01	FEED PUMP		1 x 1
108-G-02	SAND FILTER FLUSH PUMP		8 x 1
108-G-03	TAR/OIL SLOP PUMP		1 x 1
108-G-04	MUD LIQUOR PUMP		1 x 1
108-G-05	TRANSFER PUMP		3 x 0
108-G-06	EXTRACTOR MIXING PUMP		10 x 0
108-G-07	EXTRACT PUMP		2 x 0
108-G-08	REGENERATOR REFLUX PUMP		1 x 1
108-G-09	SOLVENT PUMP		1 x 1
108-G-10	TOWER FEED PUMP		1 x 1
108-G-11	TOWER BOTTOMS PUMP		1 x 1
108-G-12	PRODUCT PUMP		1 x 1

◇	SIZE AND NUMBER
○	PRESSURE PSIG
□	TEMPERATURE °F
▭	FLOW RATE DRY GAS Gals/hr
▭	FLOW RATE STEAM AND LIQUIDS lbs/hr

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE.

THIS PFD IS IDENTICAL FOR BASE CASE I AND II

Project	27-077	GEM	LURGI
Contract		00-0865	Lurgi Mineralchemical GmbH
Standard			Frankfurt (Main)
Drawn			
Checked			
Approved			
Scale			
Unit			
Base Case			
Process Flow Diagram			
Phenol Recovery			
(Phenolsolvan Unit)			
Unit 108/208			
Base Case I/II			
Drawing No.			
ME-DC-ERDA STUDY			
EF-76-C-01-2447			
Scale			
Unit			
Base Case

EQUIPMENT LIST

ITEM NO.	NAME	OPER. - SPARE
110-C-01	SYNGAS COMPRESSOR	2 + 0 (2 STAGES)
110-C-02	RECYCLE COMPRESSOR	2 + 0
110-D-01	REACTOR	4 + 0
110-E-01	SYNGAS COMPRESSOR INTERCOOLER	2 + 0
110-E-02	SYNGAS COMPRESSOR TRIM COOLER	2 + 0
110-E-03	REACTOR EFFLUENT H.E.	2 + 0
110-E-04	CONDENSER	2 + 0
110-E-05	TRIM CONDENSER	2 + 0
110-F-01	E.O. DRUM	2 + 0
110-F-02	REACTOR STEAM DRUM	2 + 0
110-F-03	EXPANSION VESSEL	2 + 0

	STEAM INLET
	PRESSURE
	TEMPERATURE
	FLOW RATE DRY GAS
	FLOW RATE STEAM
	FLOW RATE LIQUIDS

NOTES
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE

GEM 00-0865		LURGI Lurgi-Methanol-Synthese GmbH Frankfurt/Main	
PROCESS FLOW DIAGRAM METHANOL SYNTHESIS		70-GEM-6905 V-DE LURGI-STUDY 11-78-01-2447	
UNIT 110		BASE CASE I

EQUIPMENT LIST

ITEM NO.	NAME	OPER. APPROX. CAPACITY I.D. LENGTH
112-B-01	REDUCTION SYSTEM HEATER	1 x 0 Not shown on PFD
112-C-01	RECYCLE COMPRESSOR	2 x 0
112-D-01	REACTOR	3 x 0 8'-0" 12'
112-D-02	CLEAN-UP REACTOR	3 x 0 8'-0" 12'-0"
112-E-01	STEAM GENERATOR	2 x 0
112-E-02	RECYCLE BFW H.E.	2 x 0
112-E-03	FIRST FUEL/PRODUCT H.E.	1 x 0
112-E-04	PRODUCT GAS CONDENSER	1 x 0
112-E-05	PRODUCT GAS TRIM CONDENSER	1 x 0
112-E-06	SECOND FUEL/PRODUCT H.E.	1 x 0
112-E-07	PRODUCT GAS/STW H.E.	1 x 0
112-E-08	REDUCTION SYSTEM COOLER	1 x 0 Not shown on PFD
112-F-01	FIRST E.O. DRUM	1 x 0
112-F-02	SEC. E.O. DRUM	1 x 0
112-F-03	THIRD E.O. DRUM	1 x 0
112-F-04	FOURTH E.O. DRUM	1 x 0
112-F-05	REDUCTION SYSTEM E.O. DRUM	1 x 0 Not shown on PFD

◆ STREAM NUMBER
 ○ PRESSURE PSIG
 □ TEMPERATURE °F
 ▭ FLOW RATE DRY GAS Dm³/hr
 ▨ FLOW RATE STEAM AND LIQUIDS Mm³/hr

NOTE
 STEAMS WITHOUT STREAM NUMBERS ARE NOT INCLUDED IN THE OVERALL MATERIAL BALANCE.

Date: 28.07.77 Drawn by: [Signature] Checked by: [Signature] Approved by: [Signature]	GEM 00-0865	LURGI Large Mineralisierungs GmbH Frankfurt/Main
PROCESS FLOW DIAGRAM METHANATION		20-GEM-6906 MODIFIED STUDY 17-76-U-01-2447
UNIT 112 BASE CASE 1		[Signature] [Signature]

ITEM NO.	NAME	OPER. APPROX. CAPACITY T.D. (TONS/H)
113-C-01	LIGHT GAS COMPRESSOR	1 + 0
113-D-01	CO ₂ ABSORBER	1 + 0 7'-0"/81'
113-D-02	REFRIGERATOR	1 + 0 7'-0"/81'
113-E-01	REFRIGERATOR S.E.	1 + 0
113-E-02	CONDENSER	1 + 0
113-E-03	TRIM COOLER	1 + 0
113-E-04	REFRIGERATOR AND CONDENSER	1 + 0
113-E-05	REFRIGERATOR AND TRIM CONDENSER	1 + 0
113-E-06	REFRIGERATOR AND BOILER	1 + 0
113-F-01	REFRIGERATOR	1 + 0
113-F-02	REFRIGERATOR AND D.D. DRUM	1 + 0
113-F-03	REFRIGERATOR AND D.D. DRUM	1 + 0
113-F-04	REFRIGERATOR AND D.D. DRUM	1 + 0
113-G-01	SOLUTION PUMP	1 + 1
113-G-02	REFRIGERATOR REFLUX PUMP	1 + 1
114-D-01	WATER ABSORBER	1 + 0
114-D-02	TRIM REGENERATOR	1 + 0
114-E-01	TRIM COOLER	1 + 0
114-F-01	TRIM DRUM	1 + 0
114-G-01	TRIM PUMP	1 + 1

◇	STREAM NUMBER
○	PRESSURE PSID
□	TEMPERATURE °F
▭	FLOW RATE DRY GAS Gmol/hr
▭	FLOW RATE STEAM AND LIQUID lbm/hr

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE.

26777	GEM	Lurgi
	00-0865	Lurgi Process Unit
PROCESS FLOW DIAGRAM	20-GEM-6907	
CO ₂ REMOVAL (MEA WASH)	WASHER STUDY	
& SNG DRYING (TEG WASH)	EF-75-C-01-2647	
UNIT 113/114	BASE CASE I

EQUIPMENT LIST			
ITEM NO.	NAME	OPER. APPROX.	REMARKS
111-C-01	PRESSURE SWING OFFGAS RECOMPRESSION (3 STAGES)	1 + 0	
111-D-01	PRESSURE SWING ADSORBER	4 + 0	
111-F-01	PRESSURE SWING OFFGAS HOLDER	1 + 0	
151-B-01	REACTOR FEED HEATER	1 + 0	
151-C-01	RECTIFY COMPRESSOR	1 + 1	
151-D-01	REACTOR	1 + 0 3'-0"	30'
151-D-02	AMINE WASH TOWER	1 + 0	
151-D-03	NAPHTHA STABILIZER	1 + 0 3'-0"	70'
151-E-01	REACTOR EFFLUENT H.E.	1 + 0	
151-E-02	REACTOR EFFLUENT COOLER	1 + 0	
151-E-03	REACTOR EFFLUENT TRIM COOLER	1 + 0	
151-E-04	STABILIZER REBOILER	1 + 0	
151-E-05	STABILIZER CONDENSER	1 + 0	
151-E-06	STABILIZED NAPHTHA COOLER	1 + 0	
151-E-07	STABILIZED NAPHTHA TRIM COOLER	1 + 0	
151-F-01	REACTOR EFFLUENT PHASE DRUM	1 + 0	
151-F-02	AMINE WASH DRUM E.O. DRUM	1 + 0	
151-F-03	STABILIZER PHASE DRUM	1 + 0	
151-G-01	FEED PUMP	1 + 1	
151-G-02	STABILIZER REBOILER PUMP	1 + 1	
151-G-03	STABILIZER RELIEF PUMP	1 + 1	
151-G-04	AMINE CIRCULATION PUMP	1 + 1

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE

GEM 00-0865		LURGI Lurgi Mineraltechnik GmbH Frankfurt	
PROCESS FLOW DIAGRAM H ₂ RECOVERY (PRESSURE SWING) & NAPHTHA HYDROTREATING		20-GEN-6908 MODIFIED 1970/11 1970/11/24	
UNITS 111/151		BASE CASE I

ITEM NO.	NAME	QTY	APPROX. I.D. LENGTH
150-B-01	START UP & REGENERATION HEATER	1 + 0	
150-C-01	RECYCLE COMPRESSOR	1 + 0	
150-C-02	REFINERY AIR R2 COMPRESSOR	1 + 0	
150-C-03	REFINERY AIR R2 COMPRESSOR	1 + 0	
150-D-01	DWE REACTOR	2 + 0	13'-6"
150-D-02	REFINERY AIR R2 COMPRESSOR	6 + 0	30'-6"
150-E-01	DWE REACTOR FEED METHANOL CONVERTER EFFLUENT FIRST H.E.	1 + 0	
150-E-02	DWE REACTOR FEED METHANOL CONVERTER EFFLUENT SEC. H.E.	1 + 0	
150-E-03	WASTE STEAM GENERATOR	1 + 0	
150-E-04	FEED/RECYCLE FIRST H.E.	1 + 0	
150-E-05	FEED/RECYCLE SEC. H.E.	1 + 0	
150-E-06	EFFLUENT/BFW H.E.	1 + 0	
150-E-07	EFFLUENT/BFW H.E.	1 + 0	
150-E-08	REFINERY AIR R2 H.E.	1 + 0	
150-E-09	REGENERATION GAS COOLER	1 + 0	
150-F-01	EFFLUENT SETTLER	1 + 0	
150-F-02	REGENERATION RECYCLE E.O. DRUM	1 + 0	
150-G-01	DWE REACTOR FEED PUMP	1 + 1	
150-G-02	HYDROCARBON PRODUCT PUMP	1 + 1	
150-G-03	HYDROCARBON PRODUCT RECYCLE PUMP	1 + 1

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE

GEM 00-0865	LURGI Lurgi Mineralprodukte GmbH Frankfurt, Germany
PROCESS FLOW DIAGRAM METHANOL CONVERSION	20-GEM-6915 METHANOL STUDY REV. C-01-2467
UNIT 150	BASE CASE I

EQUIPMENT LIST

ITEM NO.	NAME	OPER. APPROX. SPARS. D. LENGTH TWT
152-D-01	DEETHANIZER	1 x 0 8'-0" 80'
152-D-02	STABILIZER	1 x 0 8'-0" 100'
152-E-01	DEETHANIZER REBOILER	1 x 0
152-E-02	DEETHANIZER SIDESTREAM/BOTTOMS R.E.	1 x 0
152-E-03	STABILIZER OVER CONDENSER	1 x 0
152-E-04	STABILIZER OVER PRODUCT COOLER	1 x 0
152-E-05	STABILIZER BOTTOMS REBOILER	1 x 0
152-E-06	STABILIZER SIDE REBOILER	1 x 0
152-E-07	STABILIZER BOTTOMS COOLER	1 x 0
152-E-08	STABILIZER BOTTOMS TRIM COOLER	1 x 0
152-F-01	LIGHT GAS FLASH DRUM	1 x 0
152-F-02	STABILIZER REFLUX DRUM	1 x 0
152-G-01	STABILIZER REFLUX PUMP	1 x 1

153-D-01	REACTOR SETTLER	1 x 0 8'-0" 80'
153-D-02	DEISOPHTANIZER	1 x 0 8'-0" 100'
153-D-03	DEPROPANIZER	1 x 0 8'-0" 100'
153-E-01	DEISOPHTANIZER FEED REFLUX R.E.	1 x 0
153-E-02	DEISOPHTANIZER FEED REFLUX R.E.	1 x 0
153-E-03	MIXED BUTANES COOLER	1 x 0
153-E-04	DEISOPHTANIZER SIDE REBOILER	1 x 0
153-E-05	DEISOPHTANIZER BOTTOMS REBOILER	1 x 0
153-E-06	MIXED BUTANES TRIM COOLER	1 x 0
153-E-07	DEISOPHTANIZER BOTTOMS TRIM COOLER	1 x 0
153-E-08	DEISOPHTANIZER OVER CONDENSER	1 x 0
153-E-09	DEISOPHTANIZER OVER TRIM CONDENSER	1 x 0
153-E-10	DEPROPANIZER FEED REFLUX R.E.	1 x 0
153-E-11	DEPROPANIZER FEED REFLUX R.E.	1 x 0
153-E-12	DEPROPANIZER OVER CONDENSER	1 x 0
153-E-13	DEPROPANIZER OVER PRODUCT COOLER	1 x 0
153-E-14	DEPROPANIZER BOTTOMS COOLER	1 x 0
153-E-15	DEPROPANIZER BOTTOMS TRIM COOLER	1 x 0
153-E-16	DEPROPANIZER REFLUX DRUM	1 x 0
153-F-01	DEPROPANIZER REFLUX DRUM	1 x 0
153-G-01	DEPROPANIZER FEED PUMP	1 x 1
153-G-02	DEPROPANIZER REFLUX PUMP	1 x 1
153-G-03	DEPROPANIZER REFLUX PUMP	1 x 1

NOTE
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE.

Project No.	27777 VI	GEM	Lurgi
Drawn by		00-0865	Lurgi (Microchemicals) GmbH
Checked by			Frankfurt (Main)
Scale	1/2" = 1'-0"		
Process	PROCESS FLOW DIAGRAM		ZO-GEM-6916
Unit	FRACTIONATION & ALKYLATION		W. D. C. STUDY
Drawn by	UNITS 152/153		EF-76-C-01-2447
Checked by	BASE CASE 1		

BASE CASE I
3.5 OFFSITE UNITS

Unless stated otherwise, units are shown on Block Flow Diagram ZO-GEM-6894 in Sub-Section 3.4. In addition, units such as boilers, generators and water treaters are also included in Steam/BFW Balance ZO-GEM-6910 in Sub-Section 3.6. The schematic flow diagrams (SFD) are placed at the end of this section.

3.5.1 Oxygen Production Unit 121

Purpose of the Unit is to produce the oxygen required for the gasification of the coal.

Description

The oxygen production unit is an air separation unit including the following:

- air filter
- air compressor
- air cooling
- cold box
- power recovery turbine, including electric power generator
- oxygen compressor
- nitrogen release stack
- liquid nitrogen storage tank 5,000 ft³ (for start-up, etc.)
- nitrogen evaporator, capacity 500 lb-mol/hr
- emergency liquid air evaporator
- liquid oxygen storage tank

Normal Capacity

Oxygen quantity (calculated as 100%): 14,060 lb-mol/hr
Oxygen purity: 98.06%

Design Capacity

105% of normal

Note: In addition, 600 lb-mol/hr of instrument air is supplied to the complex from this unit. (See Unit 127.)

3.5.2 Boiler Unit 122

Purpose of the Unit is to generate high pressure (HP) steam for the balance of the power requirements not provided by the various MP and LP steam generators using process waste heat.

Description

The boiler is a coal fired boiler complete with all necessary equipment required to produce 1609 psig steam, including provisions to burn the supplemental fuels and incinerate the sour offgases listed below. Efficiency assumed is 83%.

Main_Fuel

Fine coal (composition listed in Sub-Section 2.3)
Quantity 274.0 Mlb/hr
Low heating value, as received 7,893 BTU/lb
(See Unit 128 process description.)

Supplemental_Fuels

- Oil (from Gas Liquor Separation Unit 107)
Quantity 35.90 Mlb/hr
Low heating value 16,700 BTU/lb
- Tar (from Gas Liquor Separation Unit 107)
Quantity 42.85 Mlb/hr
Low heating value 15,700 BTU/lb
- Phenols (from Phenol Recovery Unit 108)
Quantity 15.26 Mlb/hr
Low heating value 7,800 BTU/lb

Incineration_Fuel

- Sour offgases (mainly from Sulphur Recovery Unit 106)
Quantity 1,592.4 Mlb/hr
Low heating value 274 BTU/lb

Note: The heat content of these gases (30 BTU/SCF) has not been included in the boiler thermal balance.

Normal Capacity 2,291.2 Mlb/hr

Design Capacity 120% of normal

3.5.3 Main Superheater Unit 123

Purpose of the Unit is to superheat the HP steam generated in the boiler for more economic use in the various HP steam turbines.

Description

The main superheater consists of all equipment necessary to produce HP steam at outlet conditions of 986°F and 1,566 psig. It is an integral part of the boiler system. Unit 122 includes the fuel requirements for the superheater. A 43.2 Mlb/hr slip stream of boiler steam is not superheated.

3.5.4 Superheater Unit 124

Purpose of the Unit is to superheat surplus medium and low steam produced in the process units.

Description

The superheater consists of all equipment necessary to superheat MP and LP steam as follows:

	<u>MP Steam</u>	<u>LP Steam</u>
Inlet Pressure, psig	571	100
Outlet Pressure, psig	550	85
Inlet Temperature, °F	483	338
Outlet Temperature, °F	748	502

In addition, each coil is controlled individually and is able to be inactive while the other is fully operating.

The superheater is designed to burn part of the oil recovered in the Gas Liquor Separation Unit 107. Efficiency assumed is 79%.

Quantity of oil	15.81 Mlb/hr
Low heating value	16,700 BTU/lb

Normal Capacity

Coil 1: MP steam superheating	765.5 Mlb/hr
Coil 2: LP steam superheating	842.7 Mlb/hr

<u>Design Capacity</u>	105% of normal
------------------------	----------------

3.5.5 Electrostatic Stack Gas Precipitator Unit 125

Purpose of the Unit is the removal of particulate matter to meet the state and federal environmental requirements for coal fired boilers. (See Sub-Section 2.6)

Description

The unit consists of all equipment required to remove the dust from Boiler Unit 122 stack gas to a level which is compatible with the downstream Stack Gas Clean Up Unit 126. The stack gas leaving Unit 126 will contain 0.1 pound of particulates per million BTU of fuel fired.

<u>Normal Capacity</u>	167.5 Mlb-mol/hr
------------------------	------------------

<u>Design Capacity</u>	120% of normal
------------------------	----------------

3.5.6 Stack Gas Clean-Up Unit 126

Purpose of the Unit is to remove SO_2 from the stack gas to meet the state and federal regulations for coal fired boilers. (See Sub-Section 2.6)

Description

The unit consists of all equipment required to remove SO_2 from the stack gas of Boiler Unit 122 to a level of 0.2 pounds SO_2 per million BTU of fuel fired. The SO_2 absorbant is a lime slurry. The clean stack gas is discharged to the atmosphere via a 300 foot stack.

<u>Normal Capacity</u>	167.5 Mlb-mol/hr, containing 36.1 lb-mol/hr SO_2
------------------------	--

<u>Design Capacity</u>	120% of normal
------------------------	----------------

3.5.7 Instrument and Plant Air Unit 127 (not shown)

Purpose of the Unit is to supply instrument air for Air Separation Unit 121 and the boiler area only, and to supply plant air for maintenance purposes. (*)

Description

The unit consists of all equipment, such as compressors and dryers, to produce instrument and plant air. The instrument air is available at 85 psig and -10°F dew point. The plant air will be available at a pressure ranging from 85 to 140 psig.

Design Capacity

Plant air	100 lb-mol/hr
Instrument air	50 lb-mol/hr

(*) Instrument air for the total complex (estimated to be 600 lb-mol/hr, normal) is supplied by the Air Separation Unit 121, from a point where the air is pressured and dried.

3.5.8 Coal Handling Unit 128

Purpose of the Unit is to prepare the as mined coal into suitable feedstocks for the gasifiers and boiler and to provide coal storage.

Technology Used

Commercial-type coal crushing and screening facilities are used to produce well-sized coal for gasification and fines for boiler firing.

The well-sized coal is in the range of 1/4" to 2", containing approximately 15% oversize and 7% undersize.

Process Description (SFD ZO-GEM-6909)

The coal is delivered by trucks from the adjacent mine area. They operate six days per week delivering approximately 32,000 tons per day. Trucks flood unload into the truck dump station having four dump pockets, each with a feeder outlet designed to accommodate two trucks simultaneously. The coal drops into a bin and is fed into the four primary crushers by four plate feeders.

Crushers reduce the size to minus 8" with a minimum generation of fines. The product of the primary crushers discharges onto two belt conveyors for weighing and conveying to the primary screening station using 8" x 20" single deck screens. The minus 2" product is fed to two belt conveyors and the plus 2" product is discharged into secondary crushers, which reduce the oversize coal to minus 2".

The minus 2" coal is then conveyed to a transfer station from where the coal is directed to the live or dead storage piles or, through a by-pass conveyor, directly to secondary screening. Normally, the conveyor to the live storage will be used. The live storage area is sufficient for about six full days of plant production. The dead storage area is sufficient to feed the plant for approximately 30 days.

Coal is reclaimed from live storage via conveyors and is transferred to the secondary screening building for final sizing into gasifier and boiler feeds. In this building, the coal is fed to the screens through variable rate vibrating feeders. The oversize from each screen falls onto two collecting conveyors for sampling and transfer to the gasifier building feed conveyors. The undersize product (minus 1/4") also falls onto two collecting belts for transfer to the fines silo.

Two belt conveyors, equipped with scales and samplers, carry the fines from the storage silo to Boiler 122.

Normal Capacity

Gasifier feed (1/4" to 2")	1,901.3 Mlb/hr
Fine Coal (minus 1/4")	376.5 Mlb/hr

3.5.9 Ash Handling Unit 129

Purpose of the Unit is to receive hot ash from the gasifier ash locks and cool the ash to a suitable temperature for continuous transport and disposal without formation of dust.

Technology Used

The ash handling system is a hydraulic sluiceway system similar to the ash system used commercially by Sasol. It has proven to be very satisfactory for handling the ash from a large number of gasifiers.

Process Description (SFD ZO-GEM-6911)

The ash from the gasifiers discharges into totally enclosed sluiceways located below grade under the gasifiers. The ash from the gasifiers is quenched by a large circulating stream of water which is recycled from the thickener clear overflow. Exhaust Fan 129-C-01 exhausts a mixture of air and water (steam) from the sluiceway. The stream of air and water is quenched with water sprays before being released to the atmosphere to reduce the water content of the stream and to eliminate any dust.

The mixture of ash and water flows by gravity to Duplex Spiral Classifier 129-Y-01 which separates the bulk of the ash from the water. Fine ash and water from the classifier flow by gravity into Classifier Sump 129-F-01 while solids from the classifier are fed to Conveyor 129-Y-02.

The water and fines from the classifier sump are pumped to Thickener 129-F-02 which is a cone bottom center feed rake thickener. The clear overflow from the thickener flows to Thickener Overflow Sump 129-F-03. From this sump, water is pumped back to the sluiceway by Thickener Overflow Sump Pump 129-G-02. The bottoms of the thickener is pumped by 129-G-01 to Filter 129-F-04. Filtrate is recycled to the thickener by 129-G-04. Solids are fed to the conveyor. The conveyor discharges the solids from the filter and from the classifier into Ash Storage Bin 129-F-05.

3.5.10 BFW Preparation Unit 131

Deaerator (131/1)

Purpose of Section 131/1 is to deaerate the boiler feed water (BFW) before steam generation.

Description

The deaerator section uses commercial-type deaeration equipment to remove dissolved gases from steam condensate and fresh water.

The dissolved gases in these streams are removed by heat supplied from Heat Exchanger 150-E-06. (See PFD ZO-GEM-6915.)

After deaeration, the BFW is pumped to the various steam generators. Stripped dissolved gases are vented to the atmosphere.

Normal Capacity

	<u>Temp., OF</u>	<u>Flow, Mlb/hr</u>
MP condensate	483	86.5
LP condensate	284	972.3
Vacuum condensate	207*	2,452.3
Methanation Unit 112		
process condensate	288	103.0
BFW make-up water	207*	2,075.9
Total	<u>221</u>	<u>5,690.0</u>

* After heating in 150-E-06

Design Capacity 120% of normal capacity

Operating Pressure 17.6 psia

Demineralizer (131/2)

Purpose of Section 131/2 is to provide the high quality water required for high pressure steam generation.

Description

Commercial-type equipment is used. The fresh water is filtered, softened and demineralized before flowing to Deaerator 131/1. (The assumed fresh water analysis is reported in Sub-Section 2.3).

<u>Normal Capacity</u>	2,125.9 Mlb/hr
<u>Design Capacity</u>	120% of normal
<u>Operating Temperature</u>	68°F (*)

(*) The difference of 20°F in the fresh water temperature listed in Sub-Section 2.3 results from the use of this fresh water stream for direct cooling of air in the Oxygen Production Unit 121.

3.5.11 Cooling Water Make-Up Preparation Unit 132

Purpose of the Unit is to treat the fresh water required to replace losses in the cooling water system.

Description

The fresh water is softened and chemically treated to become suitable for make-up to the cooling tower system. Commercial-type equipment is used. In addition to treated fresh water, the biologically treated water from Waste Water Treatment Unit 135 and the humidity of the air condensed in Air Separation Unit 121 are also used as cooling tower make-up.

The CW make-up requirement has been estimated as follows:

	<u>Percent of CW Circulation</u>
Evaporation losses	1.6
Drift	0.1
Blow-down (*)	0.5
Total	<u>2.2</u>

(*) Blow-down is based on a hardness of the softened water of 35 ppm CaCO₃.

Normal Capacity

	<u>Mlb/hr</u>
Treated fresh water	1,026.8
Treated water from Unit 135 and humidity condensed in Unit 121	<u>1,994.9</u>
Total	3,021.7

Design Capacity

Softening	200% of fresh water
Chemicals	115% of total water

3.5.12 Cooling Water Tower Unit 133

Purpose of the Unit is to supply the cooling water necessary for the steam turbine condensers and for the process/offsite unit trim coolers.

Description

Standard commercial-type cooling towers and pumps are used to provide 75°F cooling water. The cooling towers are designed for a maximum wet bulb temperature of 66°F. Make-up water is supplied to the cooling towers from Cooling Water Make-Up Preparation Unit 132.

Normal Capacity (*)

	<u>GPM</u>
Steam turbine condenser duty (temperature rise 75 to 100°F and discharge pressure 50 psig)	191,400
Trim cooler duty (temperature rise 75 to 97°F avg. and discharge pressure 64 psig)	81,725
Total	<u>273,125</u>

Design Capacity

115% of normal

(*) The CW capacity is based on air cooling to 130°F and trim cooling below 130°F.

3.5.13 Electric Power Generation Unit 134

Purpose of the Unit is to generate the electric power requirements for the plant complex.

Description

Electric power is generated from commercial-type power generators driven by steam turbines using waste steam from the process units. Two levels of steam are used to generate electric power. Generator 134-M-01 uses 48 psig, 378°F steam with Condensing Turbine 134-T-01. Generator 134-M-02 uses superheated 85 psig steam with Back Pressure Turbine 134-T-02 exhausting to 50 psig.

Electric power is generated at 6,000 Volts; stepdown transformers are provided.

Normal Capacity

	MW	Steam Flow, Mlb/hr
134-M-01	43.5	883.4
134-M-02	7.9	842.7

Design Capacity

110% of normal

3.5.14 Waste Water Treatment Unit 135

Purpose of the Unit is to treat the waste and acid waters of the total complex so that they can be used as part of the cooling water make-up requirements.

Description

The waste water treatment unit is a commercial-type, biological treatment process using air. In addition, the acid water will be neutralized using caustic soda.

Normal Capacity

<u>Composition</u>	<u>Flow, Mlb/hr</u>
Organic acids and phenols	4.8
Acetone plus C ₆ ⁺ containing some H ₂ S	2.2
Ammonia	1.3
Water	1,974.8
Total	1,983.1

Design Capacity

110% of normal capacity regarding to water flow rate
130% of normal capacity regarding to components.

3.5.15 Flare and Blow-Down Facilities 136 (not shown)

Purpose of the Unit is to provide relief for all liquid and gaseous effluents from the complex.

Description

A commercial-type flare system burns all gaseous streams coming from either process vents or safety valves. The blow-down facility consists of a holding pond to collect the cooling tower and steam system blow-down streams.

Normal Capacity

Blow down water to holding pond 809.5 Mlb/hr

Design Capacity 130% of normal

3.5.16 Storage 137 (not shown)

The following is a summary of the intermediate, product and chemical storage tank capacities which are not located within the process and offsite units:

Product Storage (15 days)

Anhydrous ammonia	1,700 ton (refrigerated)
Sulphur	1,000 ton (covered)
Gasoline	364,000 Bbl (floating roof)
Propane LPG	26,000 Bbl (pressure)
Mixed butane	37,000 Bbl (pressure)

Intermediate Storage (15 days)

Methanol	870,000 Bbl (floating roof)
Raw gasifier naphtha	22,000 Bbl (floating roof)

Chemical Storage

Diisopropyl ether	1,500 Bbl
<u>Coal Fines</u> (7 days)	8,700 T (silo)

3.5.17 Interconnecting Pipeway 138 (not shown)

Purpose of the Unit is to provide piping and racks for connection between the various process units and between the process and offsite units.

3.5.18 Refrigeration Unit 141

Purpose of the Unit is to supply the refrigeration requirements for the plant complex.

Description

An ammonia absorption refrigeration system is used. This process uses the 20 psig waste steam which cannot be utilized in another way. Two temperature levels of refrigeration are provided.

Normal Capacity

	---Duty, MMBtu/hr---	
<u>Temperature Level</u>	<u>+32°F</u>	<u>-45°F</u>
Purification Unit 105 and ammonia storage	25.5	70.5
<u>Design Capacity</u>		110% of normal

3.5.19 Gasoline Blending Unit 154

Purpose of the Unit is to blend in-line the various components continuously and automatically into a 10 RVP gasoline product.

Description

A typical commercial in-line blending operation is carried-out as follows:

- Component samples are test blended in the laboratory to determine the proper blending and additive recipe.
- Blending recipe requirements are entered into the blending control system.
- After initial checking, operation is placed on-stream and the gasoline product is sent to final storage.

Equipment consists of rundown tanks, the blending control system with valves, flow meters, analyzer and controllers, and pumps.

NOTES:
 STREAMS WITHOUT STREAM NUMBERS
 ARE NOT INCLUDED IN THE OVERALL
 MATERIAL BALANCE
 THIS SFD IS IDENTICAL FOR BASE CASE I AND II
 EXCEPT AS NOTED

GEM		Lurgi	
00-0865		20-GEM-6909	
SCHEMATIC FLOW DIAGRAM		COAL HANDLING	
UNIT 128/228		BASE CASE I/II

EQUIPMENT LIST

ITEM NO.	NAME	OPER. + SPARE
129-C-01	EXPANSION PAN	4 + 0
129-F-01	CLASSIFIER PUMP	2 + 0
129-F-02	THICKENER	4 + 0
129-F-03	THICKENER OVERFLOW DUMP	2 + 0
129-F-04	FILTER	4 + 0
129-F-05	ASH STORAGE BIN	2 + 0
129-G-01	THICKENER BOTTOMS PUMP	4 + 0
129-G-02	THICKENER OVERFLOW SUMP PUMP	2 + 2
129-G-03	CLASSIFIER SUMP PUMP	2 + 2
129-G-04	FILTRATE PUMP	4 + 0
129-Y-01	DUPLEX SPIRAL CLASSIFIER	4 + 0
129-Y-02	CONVEYER	2 + 2

NOTES:
 STREAMS WITHOUT STREAM NUMBERS ARE NOT INCLUDED IN THE OVERALL MATERIAL BALANCE.
 THIS SFD IS IDENTICAL FOR BASE CASE I AND II

Project	250777	GEM	LURGI
Client		00-0865	Large Metallurgical GmbH
Contract			Project No. (if any)
Drawn			
Checked			
Approved			
Scale			
Sheet			
SCHEMATIC FLOW DIAGRAM		20-GEM-6911	
ASH HANDLING SYSTEM		MODERNA STUDY	
(OFF SITE UNIT)		EP-76-C-D-2447	
Unit	UNIT 29.229	Base Case	1/II

BASE CASE I
3.6 UTILITY REQUIREMENTS AND STEAM/BFW BALANCE

3.6.1 Utility Requirements

Utility requirements for the individual units can be found as follows:

<u>Utility</u>	<u>Source</u>
Fuel	Type and stream number: BFD ZO-GEM-6894 (Sub-Section 3.4) Quantity and composition: Material balance sheets (Appendix A)
Power	BFD ZO-GEM-6894
Cooling water	BFD ZO-GEM-6894
Boiler feed water	Steam/BFW Balance ZO-GEM-6910 (Sub-Section 3.6)
Steam	Steam/BFW Balance ZO-GEM-6910

In addition, offsite Sub-Section 3.5 contains utility information.

3.6.2 Steam/BFW Balance

The steam and boiler feed water flows are shown on Drawing ZO-GEM-6910. The steam/electric power balance is acceptable in accordance with the study bases. Only about 5 MW(e) is surplus in comparison with 255 MW of total electric and steam power required in the plant complex. In addition, except for about 43 Mlb/hr of 20 psig steam, all steam produced is utilized.

DATE	15 4 77	NAME	GEM	ORG	LURGE
TIME	00-0865				Large Atmospheric Shell Skid
NO. OF CARS					By Air
STEAM/BFW BALANCE					ZO-GEM-6910
COAL TO GASOLINE CONV.					MRDC/ERDA STUD
VIA METHANOL					EF-76-C-01-244
BASE CASE I					By Air
					Super on load, etc.

BASE CASE I
3.7 TRAIN PHILOSOPHY

The train philosophy for the various units was selected on the basis of providing a plant complex having an onstream factor of 92%. Consequently, parallel trains have been used for many units to give flexibility in both operation and maintenance. Also, because of the large size of the complex, some units must be divided into parallel trains. In the methanol conversion and hydrocarbon processing area, however, intermediate storage is used to provide this flexibility.

In addition, all normally operating pumps, reciprocating compressors and conveyors are provided with an 100% spare.

The following is a summary of the train philosophy used for each unit:

Gasification Unit 101

Two parallel trains, each containing 14 Lurgi gasifiers, are used. Each gasifier includes a wash cooler, waste heat boiler and other associated equipment.

Only 25 gasifiers will be required to produce the design synthesis gas quantity. The other three spare gasifiers can be withdrawn from service for maintenance.

Raw Gas Shift Unit 102

Two parallel trains are provided because of reactor size limitations and to provide operating flexibility.

Raw Gas Cooling Unit 103

Three parallel trains are provided due to the size limitation of the heat exchangers.

Shifted Gas Cooling Unit 104

One train is provided since the heat exchange equipment is small and a centrifugal booster compressor is provided.

Gas Purification Unit 105

The absorber and regenerator section is divided into four parallel trains to allow towers to be shop fabricated. All the other sections of the unit are divided into two parallel trains for operating flexibility.

Sulphur Recovery Unit 106

The absorber section is divided into four parallel trains due to the large size of the absorbers. In addition, because of the height requirements of the absorbers, each train will contain two absorbers in series. All other sections of this unit are divided into two parallel trains for operating flexibility.

Gas Liquor Separation Unit 107

The dusty gas liquor separation equipment is divided into three trains to provide operating flexibility and to keep the equipment reasonably sized.

Phenol Recovery Unit 108

The phenol extraction section is divided into two parallel trains for operating flexibility and for reasonably sized equipment.

Ammonia Recovery Unit 109

The total stripper and associated equipment is divided into two parallel trains because of size limitations.

Methanol Synthesis Unit 110

Because of manufacturing size limitations, this unit is divided into four parallel reactor trains. Also, two feed/recycle centrifugal compressors are provided for operating flexibility.

H₂ Recovery Unit 111

One train is provided.

Methanation Unit 112

Two parallel reactor trains are provided for operating flexibility.

CO₂ Removal Unit 113

One train is provided.

SNG Drying Unit 114

One train is provided.

Methanol Conversion Unit 150
Naphtha Hydrotreating Unit 151
Fractionation Unit 152
Alkylation Unit 153

All are single train units. They are decoupled from the gasification and methanol synthesis sections by 15 days of tank capacity for the crude methanol and raw naphtha. In addition, Unit 150 has the following reactor configurations for efficient operability:

<u>Reactors</u>	<u>Design</u>
First-stage DME	2 in parallel for 1 year regeneration
Second-stage M-Gasoline	6 in parallel for 14 day cyclic regeneration

Offsite Units

The offsite units are designed for a minimum of two parallel trains. In cases where commercial-sized equipment is too large with two trains, the number of trains has been increased.

SECTION 4
BASE CASE II
FISCHER-TROPSCH TECHNOLOGY

BASE CASE II
4.1 MATERIAL BALANCE

The overall material balance is shown below:

<u>Input</u>	<u>Mlb/hr</u>
Coal, as mined	2,316.0 ⁽¹⁾
Air	7,309.6
Water	3,319.9
Other	7.9
	<hr/> 12,953.4
<u>Output</u>	<u>Mlb/hr</u>
Ash	147.8
Products, including alcohols	519.7 ⁽²⁾
Byproducts	13.7
Blowdown water	837.9
Stack and evaporation losses	11,428.4
Other	5.9
	<hr/> 12,953.4

(1) 27,792 T/SD

(2) 44,950 FOE Bbl/SD @ 6.0 MM Btu/FOE Bbl

Overall plant consumptions per FOE barrel of product are:

Coal:	0.618 T
Water:	5.05 Bbl
Air:	1.95 T

Stack and evaporation losses amount to about 3.0 tons per barrel of product.

Appendix B contains the complete, detailed material balance for Base Case II.

BASE CASE II
4.2 PRODUCT YIELDS AND QUALITY

Stream numbers are given below for reference to the material balance sheets in Appendix B and process diagrams in Sub-Section 4.4.

4.2.1 SNG (49)

<u>Quantity</u>	19,023.3 lb-mol/hr
	315.4 Mlb/hr
	173.3 MMSCF/SD

Composition (Only compounds greater than 0.1% are listed; see Appendix B for detailed composition.)

	<u>Percent</u>
Hydrogen	3.8
Methane	89.7
Ethene	1.0
Ethane	2.3
Propene	1.0
Propane	0.1
Carbon dioxide	0.5
Inerts (N ₂ + Ar)	1.6
	100.0

Other

Heat of combustion (HHV)	1,003 Btu/SCF
Carbon monoxide (0.1% max.)	0.07%
Water	0.01%
Sulfur	None

Compatibility Indexes (versus pure methane)

<u>Index</u>	<u>Calculated</u>	<u>Preferable</u>	<u>Objectionable</u>
Lifting, I _l	0.99	under 1.0	above 1.06
Flash-back, I _f	1.06	under 1.15	above 1.2
Yellow-tip, I _y	0.97	above 1.0	under 0.8

The SNG product is of satisfactory quality and is interchangeable with pure methane. The olefin nature of the Fischer-Tropsch light gases, however, gives rise to a yellow-tip index value slightly below the preferable value. If the olefins prove troublesome in actual tests, an expensive hydrogenation step for olefin saturation could be added.

4.2.2 Gasoline (50)

Quantity

13,580 Bbl/SD
140.9 Mlb/hr

Blending

<u>Component</u>	<u>Wt. %</u>
Mixed butanes	2.7
Light hydrocarbons	2.9
Alkylate	2.7
Light polymer gasoline	16.3
Hydrogenated heavy polymer gasoline	14.3
C5/C6 isomerate	25.1
Reformate	23.8
Hydrotreated gasifier naphtha	11.0
	<u>100.0</u>

Properties

Estimated properties and a comparison with unleaded gasoline specifications are presented in Table 4.2.1.

This gasoline meets all specifications. Compared to typical present-day gasolines, it is slightly lower in gravity and contains fewer aromatics.

Heavy polymer gasoline hydrogenation has been added in order to meet the olefin target of 20 vol. % max. Without hydrogenation, the following are estimated:

$$\begin{aligned}(\text{Research} + \text{Motor})/2 &= 88.5 \\ \text{Olefin, vol. \%} &= 28\end{aligned}$$

Experience with high olefinic gasolines is limited. Consequently, marketing such a gasoline would require extensive testing and, perhaps, additive package reformulation. While the olefin concentration of 20 vol. % is considered tolerable, the quality of the F-T gasoline is actually poorer than the quality of the Base Case I gasoline having only 11 vol. % olefins.

TABLE 4.2.1

COMPARISON OF PRINCIPAL UNLEADED GASOLINE SPECIFICATIONS
WITH ESTIMATED FISCHER-TROPSCH GASOLINE PROPERTIES -
CASE II

	<u>Estimated F-T Unleaded Gasoline Properties</u>	<u>Specifications</u>
Gravity, °API	67.2	
Octane Numbers		
Research	91	
Motor	83	82 min.
(Research + Motor)/2	87	87 min.
Volatility		
Reid Vapor Pressure, lb	10.0	
Distillation, °F		
IBP	86	
10%	108	158 max.
30%	137	
50%	186	170/250 min./max.
70%	249	
90%	335	374 max.
EP	420	437 max.
V/L Ratio (=20), °F	@127	@140 max.
Sulfur, wt. %	Nil	0.10 max.
Composition, vol. %		
Paraffins	60	
Olefins	20	20 max. (target)
Naphthenes	3	
Aromatics	17	
Molecular Weight	93	

4.2.3 Diesel Fuel (117)

Quantity

2,307 Bbl/SD
25.4 Mlb/hr

Properties

Estimated properties and a comparison with diesel fuel specifications are shown in Table 4.2.2.

This product could be marketed as a premium diesel oil, Grade No. 1-D, for engine service requiring frequent speed and load changes. Its relatively high estimated cloud point, however, could preclude it being marketed in cold weather regions.

4.2.4 Heavy Fuel Oil (118)

Quantity

622 Bbl/SD
7.4 Mlb/hr

Properties

Estimated properties and a comparison with gas turbine fuel oil specifications are shown in Table 4.2.3.

Because of the non-existent sulfur and metals contents, this product could be marketed as a premium gas turbine fuel.

4.2.5 Alcohols (156)

Quantity

1,825 B/SD
21.2 Mlb/hr

Composition

Composition of the mixed alcohol product is:

	<u>Wt. %</u>
Ethanol	59
I-propanol	14
N-propanol	14
2-butanol	4
I-butanol	1
N-butanol	5
2-pentanol	1
N-pentanol	1
C ₆ plus	1
	<hr/> 100

The alcohol product is essentially free of acids, aldehydes, ketones and water.

TABLE 4.2.2

COMPARISON OF PRINCIPAL DIESEL FUEL SPECIFICATIONS WITH
ESTIMATED FISCHER-TROPSCH DIESEL FUEL PROPERTIES - CASE II

	<u>DIESEL FUEL-ASTM D975(1)</u>			<u>Estimated F-T Diesel Fuel Properties</u>
	<u>Grade No. 1-D</u>	<u>Grade No. 2-D</u>	<u>Other</u>	
Gravity, °API	-	-		56
Cetane Number, min. (D613 or D976)	40	40	45(2)	>50
Volatility				
90% min./max., °F (D86)	-/550	540/640(5)		520
Flash Point, °F min. (D93)	100	125		125
Cloud Point, °F (D2500)	(3)	(3)		10-20
Viscosity, Kin. @ 100°F min./max. (D445)	1.4/2.5	2.0/4.3		2
Sulfur, wt. % max. (D129)	0.50	0.50	(4)	None
Corrosion, max. (D130)	No. 3	No. 3		No. 1
Carbon Residue on 10% Resid., max. (D524)	0.15	0.35		Trace

-
- (1) 1976 Annual Book of ASTM Standards.
 (2) Various engine builders' requirements.
 (3) Varies with marketing region; maps included in ASTM D975.
 (4) Varies in individual states, some as low as 0.3%.
 (5) 540/680 90% min./max. proposed but currently not accepted by industry.

TABLE 4.2.3

COMPARISON OF PRINCIPAL FUEL OIL AND GAS TURBINE FUEL OIL SPECIFICATIONS WITH
ESTIMATED FISCHER-TROPSCH HEAVY FUEL OIL PROPERTIES - CASE II

<u>Fuel Oil</u>	<u>Fuel Oil No. 5 (Light) ASTM D396*</u>	<u>Gas Turbine F.D. No. 3 - GT ASTM D2880*</u>	<u>Estimated F-T Hvy. Fuel Oil Properties</u>
Gravity, °API	-	-	41
Volatility			
Flash Point, °F min. (D93)	130 or Legal	130 or Legal	150
Distillation	-	-	-
Ash, wt. % max. (D482)	0.10	0.03	Trace
Viscosity, SSU at 100°F, min. (D445)	150	45	Satisfactory
Sulfur, wt. % (D129)	Legal	-	None
Metals, ppm (wt.) max.			
Vanadium	-	2	0
Sodium + Potassium	-	5	Trace
Calcium	-	10	0
Lead	-	5	0

*1976 Annual Book of ASTM Standards.

4.2.6 Butane LPG (139)

Quantity

146 Bbl/SD
1.2 Mlb/hr

Properties

Estimated properties and a comparison with LPG specifications are shown in Table 4.2.4.

This product is a satisfactory commercial butane fuel.

4.2.7 Propane LPG (138)

Quantity

1,107 Bbl/SD
8.2 Mlb/hr

Properties

Estimated properties and a comparison with LPG specifications are shown in Table 4.2.4.

This product is a satisfactory commercial propane fuel.

4.2.8 Byproducts

Sulfur (29)

Quantity
Impurities

61 T/SD
0.5 wt. %

Anhydrous Ammonia (17)

Quantity
Impurities (mainly water)
Grade

103 T/SD
0.5 wt. % max.
agricultural

Excess Power (6,000 volts)

3.31 ME (e)

TABLE 4.2.4

COMPARISON OF PRINCIPAL LPG SPECIFICATIONS WITH
ESTIMATED FISCHER-TROPSCH LPG PROPERTIES - CASE II

	Commercial Propane <u>ASTM D1835*</u>	Est. F-T C ₃ LPG <u>Properties</u>	Commercial Butane <u>ASTM D1835*</u>	Est. F-T C ₄ LPG <u>Properties</u>
Vapor Pressure at 100°F, max., psig (D1267)	208	178	70	37
Volatile Residue				
Butane and Heavier, max. vol. % (D2163)	2.5	1	-	-
Pentane and Heavier, max. vol. % (D2163)	-	-	2.0	1
Corrosion, max. (D1838)	No. 1	No. 1	No. 1	No. 1
Sulfur, grains (D2784)	15	None	15	None

*1976 Annual Book of ASTM Standards.

BASE CASE II

4.3 THERMAL EFFICIENCY

Table 4.3.1 is a summary of the overall plant thermal efficiency, which is defined as the combustion heat of the products, byproducts and excess power divided by the combustion heat of the total coal feed. The efficiency has been calculated using both the high and low heating values. The HHV thermal efficiency is 57.6% and the LHV is 54.6%.

As a comparison to the efficiencies shown in Table 4.3.1, the Sasol II thermal efficiency (HHV) is expected to be 35 - 38% during the early phases of operation and is expected to increase to about 60% as the demand for the coproduced gases increases. (16) The lower efficiency is based on reforming the light gases into additional synthesis gas, and the higher efficiency is based on marketing these gases. It is understood that a "town gas" consisting of a mixture of gases will be sold locally. A methanation step to convert these gases into a high Btu SNG will not be utilized. If methanation were installed, the Sasol II thermal efficiency would approach the values shown for Base Case II.

TABLE 4.3.1

FISCHER-TROPSCH CASE II - THERMAL EFFICIENCY

	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 ⁽¹⁾	-
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
C3 LPG	1,107Bbl	3.815MMBtu/Bbl	176	0.9	3.514MMBtu/Bbl	162	0.9
C4 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
Sub-Total			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 ⁽²⁾	0.1
Total			11,351	57.6		10,346	54.6

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

(2) Direct thermal conversion used.

BASE CASE II
4.4 PROCESS DESCRIPTIONS

The discussions below follow the block flow diagram (BFD) or the process flow diagram (PFD) as noted. These diagrams are placed at the end of the discussion. The gasification, raw gas treating and byproduct recovery units, however, are identical to those in Base Case I; consequently, their PFD's are found in Sub-Section 3.4. Detailed compositions and flows of the numbered streams are found on the material balance sheets in Appendix B.

4.4.1 General (BFD ZO-GEM-6988)

Sized coal is received from the offsite Coal Handling Unit 228 (see Sub-Section 4.5), and is gasified in Gasification Unit 201. A portion of the raw gas generated is converted in Raw Gas Shift Unit 202 to meet the H_2/CO ratio requirements of F-T Synthesis Unit 250. The raw gas and shifted gas are cooled in Raw Gas Cooling Unit 203 and Shifted Gas Cooling Unit 204, respectively, before flowing to Gas Purification Unit 205. This unit removes the H_2S , CO_2 and naphtha from the raw gas. The H_2S contained in the purification unit offgas is converted to sulphur in Sulphur Recovery Unit 206. The purified gas (synthesis gas) is sent to F-T Synthesis Unit 250 where it is converted to hydrocarbon products.

The gas liquor condensed in Units 201, 202, 203, 204, and 205 flows to Gas Liquor Separation Unit 207 in which dusty tar and oil are recovered. The gas liquor is further processed in Phenol Recovery Unit 208 and Ammonia Recovery Unit 209 before flowing to Waste Water Treatment Unit 235.

The products from the F-T synthesis are upgraded and recovered in the following units: F-T Product Fractionation Unit 252, F-T Product Hydrotreating Unit 253, Hydrotreated Product Fractionation Unit 254, Catalytic Reforming Unit 255, C_5/C_6 Isomerization Unit 256, Catalytic Polymerization Unit 257, HF Alkylation Unit 258, Poly Gasoline Hydrogenation Unit 259, Light Ends Recovery Unit 260 and Alcohol Recovery Unit 262. The final products produced in these units are gasoline, diesel fuel oil, heavy fuel oil, n-butane LPG, propane LPG and alcohols.

In addition to the liquid products, SNG is produced in the SNG-upgrading train from the light gases produced in F-T Synthesis Unit 250. This train consists of Hydrocarbon Recovery Unit 210, Methanation Unit 212, where unreacted CO , H_2 and hydrocarbons are converted to methane, CO_2 Removal Unit 213, and SNG Drying and Compression Unit 214.

The hydrogen required for hydrotreating the F-T synthesis products and the naphtha from gasification is produced from hydrogen containing offgases in Hydrogen Recovery Unit 211 and Hydrogen Purification Unit 261.

4.4.2 Gasification Area (BFD ZO-GEM-6988)

Because the study basis calls for the same synthesis gas production in both base cases, the units in the coal gasification area, including gas treating and product recovery, are identical.

<u>Unit</u>	<u>PFD ZO-GEM</u>
201-Gasification (In Base Case I, Lock Gas Compressor 101-C-41 recompresses the syngas compressor losses. In Base Case II, these losses have been neglected.)	6895 & 6896
202-Raw Gas Shift	6897
203-Raw Gas Cooling	6898
204-Shifted Gas Cooling	6899
205-Gas Purification	6900
206-Sulfur Recovery	6901
207-Gas Liquor Separation	6902
208-Phenol Recovery	6903
209-Ammonia Recovery	6904

Consequently, descriptions of these units are not repeated in this sub-section.

4.4.3 Hydrocarbon Recovery Unit 210 (PFD ZO-GEM-6973)

Purpose of the Unit is to remove and recover heavy hydrocarbons from the Fischer-Tropsch synthesis purge gas to meet methanation feedstock specifications.

Technology Used

The commercial process used is a low temperature heptane wash. This process is conventionally used to handle a gas containing CO₂ because the operating temperature is high enough to prevent freezing of the CO₂.

Process Description

The feedgas is cooled against the product gas and by refrigeration in 210-E-01, 02, and 03 to condense the bulk of the higher hydrocarbons before being charged to Wash Absorber 210-D-01. To prevent freezing of water, a small alcohol stream is injected into the feedgas. In the absorber, the feed gas flows counter-currently to the heptane solvent which is pumped from Heptane Regenerator 210-D-02 through Heat Exchanger 210-E-05 and Chiller 210-E-04 to the top of the absorber.

The rich heptane leaving the absorber is fed to the heptane regenerator in which the absorbed hydrocarbons are removed by fractionation. The regenerator overhead product vapor is compressed by 210-C-01, cooled and combined with the liquid stream which is condensed from the feed gas in front of the absorber. The combined stream is sent to F-T Product Fractionation Unit 252.

The absorber overhead product gas flows to Methanation Unit 212.

4.4.4 H₂ Recovery Unit 211 (PFD ZO-GEM-6974)

Purpose of the Unit is to recover hydrogen for use in the plant hydrogenation/hydrotreating units. The feedstock is a slip stream of the product gas from hydrocarbon recovery.

Technology Used

Because of the relatively high CO₂ content of the feedgas, the Pressure Swing Process of Union Carbide Corp. is used.

Process Description

The unit consists of four parallel pressure-swing absorbers, 211-D-01, with one onstream and the other three undergoing regeneration. During operation, essentially all of the non-hydrogen components and a portion of the hydrogen are absorbed. The remaining hydrogen leaves the absorber at a purity of over 98%.

Regeneration is by pressure reduction in 3-stages with the final pressure being nearly atmospheric. The system is controlled automatically by motor valves. An absorption/regeneration cycle requires several minutes.

The desorbed gases, called pressure-swing offgases, are recompressed by Compressor 211-C-01 and returned to the SNG upgrading train as feed to Methanation Unit 212. Gas Holder 211-F-01, on the suction side of the compressor, is required for buffering the non-static, pressure-swing

offgas flow from the regenerating absorbers.

For clarity purposes, the hydrogen-source streams and the hydrogen-feed streams are shown as part of the PFD for Units 211 and 251. Hydrogen compressors 211-C-03 and 04 are also included.

4.4.5 Methanation Unit 212 (PFD-ZO-GEM-6975)

Purpose of the Unit is the production of SNG meeting the specifications for interchangeability with existing pipeline gases. In addition, the CO content must be lowered to less than 0.1%.

Technology Used

The Lurgi Methanation Process is used.
(For addition comments, see Paragraph 3.4.13.)

Process Description

The feedgas is the purge gas from F-T Synthesis Unit 250, from which the bulk of the higher hydrocarbons has been removed in Hydrocarbon Recovery Unit 210 in order to prevent fouling of the methanation catalyst.

The methanation unit consists of two down-flow, fixed-bed reactors in series. The feedgas is heated in 212-E-03, 04, and 05 against the second reactor, 212-D-02, product. Steam is injected in the feedgas, and the gas is also diluted with a portion of the methanated gas from the first reactor, 212-D-01. The dilution gas is compressed by Recycle Compressor 212-C-01, after having been cooled by 212-E-01 and 02. MP steam is generated in 212-E-01.

The net reactor effluent from the first reactor flows to the second, or clean-up, reactor in which final methanation takes place. The clean-up reactor effluent is heat exchanged against the gas feed and cooled in Exchangers 212-E-06 and 07.

Process water condensed from the reactor effluent stream is collected and routed to BFW preparation Unit 231.

The methanated gas leaves the unit for further treating CO₂ Removal Unit 213.

4.4.6 CO₂ Removal Unit 213 (PFD ZO-GEM-6976)

Purpose of the Unit is to lower the CO₂ content in the SNG to about 0.5%, thereby improving its quality.

Technology Used

The Amine Guard Process of Union Carbide Corp. is used. (For additional comments, see Paragraph 3.4.14.)

Process Description

The Methanation Unit 212 product gas and the light gases from F-T Product Fractionation Unit 252 are charged to Absorber 213-D-01, in which the CO₂ is removed by a countercurrent stream of lean MEA solution. In the top of the absorber, the gas is scrubbed with a small flow of BFW to recover trace amounts of MEA solution. The purified gas then flows to SNG Drying and Compression Unit 214.

The absorber bottoms product (rich MEA solution) is expanded after heat exchange in 213-E-01 into MEA Regenerator 213-D-02, where the CO₂ and other co-absorbed compounds are stripped from the MEA solution. The steam in the regenerator overhead is condensed in Condensers 213-E-04 and 05, and the remaining offgases are sent to Boiler Unit 222 for incineration. The steam condensate is refluxed to the top of the regeneration. The lean solution from the bottom of the regenerator is pumped back to Absorber 213-D-01 after being cooled in 213-E-01, 02 and 03.

4.4.7 SNG Drying and Compression Unit 214 (PFD ZO-GEM-6976)

Purpose of the Unit is to dry the wet SNG to a water content of 4 lb/MMSCF and to compress the dried SNG for delivery into the SNG pipeline system.

Technology Used

The TEG Wash Process, a commercial process for drying natural gas, is used.

Process Description

The drying agent is a mixture of triethylene glycol (TEG) and other ethylene glycols.

The SNG is contacted countercurrently with lean TEG in Water Absorber 214-D-01. The water-rich TEG flows from the absorber to TEG Regenerator 214-D-02 which is heated by fuel gas. The steam stripped from the TEG is discharged to the atmosphere. The small reflux stream required is condensed within the vent tube by air cooling.

The regenerated TEG is stored in 214-F-01, where it is cooled by the rich TEG before being pumped into the top of Water Absorber 214-D-01.

The dried SNG is compressed by 214-C-01 for delivery into the SNG pipeline system.

4.4.8 Fischer-Tropsch Synthesis Unit 250 (PFD ZO-GEM-6977)

Purpose of the Unit is to synthesize hydrocarbons by reacting hydrogen and carbon monoxide.

Technology Used

Since the early 20's when Fischer and Tropsch produced higher hydrocarbons from carbon monoxide and hydrogen, the amount of research and development in the field of carbon monoxide hydrogenation has been overwhelming. Today, this chemistry can produce a full range of petrochemical and oil products, including methane (methanation), methanol, and liquid hydrocarbons and waxes (Fischer-Tropsch). Before 1940, commercial Fischer-Tropsch plants were operating in Germany, France and Japan. Low price crude oil after World War II precluded any further commercial use of F-T technology, the exception being Sasol in Sasolburg, South Africa. This plant, completed in 1955, is still in operation (Sasol I) and a large expansion is currently in progress (Sasol II). It is the Sasol technology, as reported in the literature, which is the technical basis for Unit 250 development.

Although a typical F-T product is a complicated mixture, the following overall equations illustrate the process:

These equations do not indicate the reaction mechanism, but they do show the stoichiometric relationship between reactants and products. Except for the water gas shift reaction, all reactions are highly exothermic. The F-T product includes paraffins, mono-olefins, aromatics, alcohols, aldehydes, ketones and fatty acids with carbon numbers from 1 through 35+. Small amounts of diolefins and esters can also be produced. Branching does occur but it is predominantly a single-methyl branched structure. Yield selectivity and composition is highly dependent upon the catalyst, reaction conditions and reactor type.

Commercial catalysts include cobalt (fixed-bed) and iron (fixed-and fluid-bed.) Both are promoted for improved activity and selectivity. Operating conditions range from 200 to 325°C and from atmospheric to 25 atm pressure, depending upon products desired, catalyst and reactor design. For example, conditions for the Sasol I medium pressure units are reported to be (15):

<u>Type</u>	<u>Fixed-Bed</u>	<u>Fluid-Bed</u>
Catalyst Base	Fe	Fe
Temperature, °C	220-255	320-330
Pressure, psig	360	330

Although iron is the base for both units, catalyst preparation and formulation are entirely different and very specific for each unit type.

The principal problem in designing the reactor is heat removal. Fixed-bed reactor designs are some form of a heat exchanger with catalyst being cooled by boiling water or circulating oil. Fluid-bed reactors include internal tube bundles for reaction heat removal.

As stated previously, Unit 250 is based on Sasol-type fluid-bed technology, derived mainly from references cited which include both Sasol articles and those reporting on Sasol operation. The projected commercial fluid-bed process design conditions and yields are as follows:

1. Feed Ratio $\text{H}_2 / (2\text{CO} + 3\text{CO}_2) = 1.03$ Ratio is critical. (9) (12)

2. Basic yields (7) (8)

	<u>C atoms</u>	<u>H atoms*</u>	<u>H₂O atoms*</u>
C ₁	10	40.00	-
C ₂ =	4	8.00	-
C ₂	6	18.00	-
C ₃ =	12	24.00	-
C ₃	2	5.33	-
C ₄ =	8	16.00	-
C ₄	1	2.50	-
C ₅ plus	8	16.50	-
light oil	35	69.30	1.25
heavy oil	7	13.84	0.08
alcohols	6	11.28	2.70
acids	1	0.21	0.90
	<u>100</u>	<u>224.96</u>	<u>4.93</u>

*Estimated by MRDC.

3. Yields adjusted by MRDC to obtain 85.0% conversion of (CO + H₂) (8)

CO: 17,806.5 lb-mol/hr reacted = 96.25% conversion
H₂: 37,857.6 lb-mol/hr reacted = 80.57% conversion
55,664.1 lb-mol/hr reacted = 85.0% conversion

C ₁	28,843.6 lb/hr
C ₂ =	10,087
C ₂	16,219
C ₃ =	30,262
C ₃	5,285
C ₄ =	20,175
C ₄	2,612
C ₅ plus	21,196
light oil	89,397
heavy oil	17,181
alcohols	23,744
acids	5,110
	<u>270,111.6</u>

A detailed yield breakdown is presented in Table 4.4.1.

TABLE 4.4.1

FISCHER-TROPSCH CONVERSION UNIT 250
PRODUCT COMPOSITION

Component	MW	Syn Gas Feed, lb/hr	From Feed Gas, lb/hr	From Reaction, lb/hr	From C ₅ Plus (2), lb/hr	From Light Oil (3), lb/hr	From Heavy Oil (4), lb/hr	Total, lb/hr	Total, lb-mol/hr
N ₂	28.02	8,658.2	8,658.2					8,658.2	309.0
CO ₂ (1)	44.01	127,629.0	136,871.1					136,871.1	3,110.0
CO(1)	28.01	518,185.0	13,542.8					13,542.8	483.5
H ₂ (1)	2.016	94,721.8	18,824.2					18,824.2	9,337.4
C ₁	16.04	199,858.4	199,858.4	28,843.6				228,702.0	14,258.2
C ₂ -	28.05	589.0	589.0	10,087.0				10,676.0	380.6
C ₂	30.07	8,329.4	8,329.4	16,219.0		40.2		24,588.6	817.7
C ₃ -	42.08			30,262.0		1,099.6		31,361.6	745.3
C ₃	44.09			5,285.0		201.2		5,486.2	124.4
C ₄ -	56.10			19,036.0		2,145.5		21,181.5	377.6
I C ₄ -	56.10			1,139.0		134.1		1,273.1	22.7
N C ₄	58.12			2,612.0		402.3		3,014.3	51.9
1-C ₅ -	70.13				7,690.0	3,754.7		11,444.7	163.2
IC ₅	70.13				3,204.0	1,564.4		4,768.4	68.0
NC ₅	72.15			(2)	1,922.0	938.7		2,860.7	39.6
1-C ₆ -	84.16			21,196.0	3,684.0	4,337.4		8,021.4	95.3
I C ₆ -	84.16				2,210.0	2,602.4		4,812.4	57.2
NC ₆	86.17				1,473.0	1,734.9		3,207.9	37.2
200	97				1,013.0	8,824.9		9,837.9	101.4
250	110					15,329.4		15,329.4	139.4
300	127/122					8,330.0	527.7	8,857.7	69.9
350	144			(3)		13,345.0		13,345.0	92.7
400	165/163			89,397.0		6,740.4	627.4	7,367.8	44.6
450	187					9,812.9		9,812.9	52.5
500	211/210					2,977.5	1,627.1	4,604.6	21.8
550	238/232					5,081.5	993.7	6,075.2	25.7
600	280						2,185.0	2,185.0	8.4
650	288						1,707.9	1,707.9	5.9
700	315			(4)			2,944.1	2,944.1	9.3
750	350			17,181.0			2,199.8	2,199.8	6.3
800	380						1,424.7	1,424.7	3.7
850	409						1,082.7	1,082.7	2.6
900	440						366.7	366.7	0.3
1,000 Plus	510						1,494.2	1,494.2	2.9
Alcohols	51.34	42.0	42.0	23,744.0				23,786.0	463.3
Acids	65.26			5,110.0				5,110.0	78.3
H ₂ O(1)	18.016			301,186.1				301,186.1	16,717.7
		958,012.8	386,715.1	571,297.7	21,196.0	89,397.0	17,181.0	958,012.8	48,326.0
			(958,012.8)						

(1) Yields reflect shift equation equilibrium.

4. Light and heavy oil carbon atom distribution and distillation, alcohol yield breakdown, and acid yield breakdown are from Reference 7.
5. Reactor design parameters used are:
 - Heat of reaction (MRDC estimate): 4,550 Btu/lb total product
 - Fresh feed charge: 207 MNm³/hr per reactor.

The literature, (10) & (14), refers to a second generation fluid-bed reactor being from 2 to 3 times the size of Sasol I. Reference 8 used a Sasol I reactor scale-up of 2 in their study. Using Reference 8 as the basis, the above charge rate per reactor represents a scale-up of 2.1. Five reactors are employed with one being a spare.

- Recycle mole ratio (15): 2/1
- Catalyst fill (MRDC estimate from (9) (12)):
335 metric tons per reactor.
- Operating conditions (7)
 - vapor feed temperature: 160°C
 - catalyst and vapor outlet temperature: 340°C
 - vapor outlet pressure: 300 psia
- Catalyst life (7) (8): 50 days

Process Description

The synthesis gas from Purification Unit 205 and the recycle gas are compressed together in Compressor 250-C-01, heat exchanged against a portion of the reactor effluent in Exchanger 250-E-01 and charged to the fluid-bed Reactor 250-D-01. After mixing with the circulating hot Fe catalyst, reaction takes place as the mixture flows up the reactor through tube bundles in which oil is pumped (250-G-01) for heat removal (temperature control). At the top of the reactor, the mixture enters a large vessel in which cyclones are installed for iron/vapor separation. The hot oil is circulated to Steam Generator 250-E-02 where 200 psig steam is produced.

Because of the expected fouling nature of the reactor effluent, it is washed and cooled from 645°F to 340°F in Effluent Wash Tower 250-D-02 by a recycle stream from the bottoms of Vapor Cold Wash Tower 250-D-03. A relatively clean, hot stream is withdrawn from Effluent Wash Tower 250-D-02 for cooling. Its heat is used for heating the reactor vapor feed (250-E-01), boiler feed water heating (250-E-06) and LP steam generation (250-E-07 and 08). A hot heavy oil is withdrawn from the bottom and sent to Fractionation Unit 252.

The overhead vapor is cooled in BFW Exchanger 250-E-03 and Condensers 250-E-04 and 05. It is then water washed in Wash Tower 250-D-03. The overhead vapor is split into recycle and purge gas streams. The latter is sent to Hydrocarbon Recovery Unit 210. The condensed liquid is also split into 2 streams: cold recycle to Tower 250-E-02 and light oil product to Wash Tower 250-D-04 for final water washing. The light oil is sent to Fractionation Unit 252. The used wash water, containing the alcohols and acids from the Fischer-Tropsch reaction, flows to Alcohol Recovery Unit 262.

4.4.9 Naphtha Hydrotreating Unit 251 (ZO-GEM-6974)

Purpose of the Unit is yield a clean, saturated stabilized naphtha for gasoline pool blending from the raw gasifier naphtha.

Technology Used

See Paragraph 3.4.17.

Process Description

Identical to Unit 151. See Paragraph 3.4.17.

4.4.10 F-T Product Fractionation Unit 252 (ZO-GEM-6978)

Purpose of the Unit is to separate the F-T reactor product into light gases, feedstocks for the polymerization and hydrogenation units and a 850+ residuum for boiler fuel.

Technology Used

Because of the uncertainty as to the effect of steam on the remaining alcohols and acids, a 4-tower, dry system is used.

Process Description

The heavy oil is heated in Furnace 252-B-01, flashed in Drum 252-F-01, and charged to Vacuum Tower 252-D-01. Vacuum is maintained by 252-C-01. The tower has an upper pump-around system. The overhead vapors and excess pump-around liquid are condensed/cooled in Condenser 252-E-03 and pumped (252-G-15) to the top of Absorber/Deethanizer 252-D-02. Vacuum residue (bottoms) is pumped (252-G-03) to Boiler Unit 222.

Feed to Absorber Deethanizer 252-D-02 consists of (1) light oil from Unit 250, (2) hydrocarbons recovered in Unit 210, (3) vacuum tower overhead, and (4) lean oil from Naphtha Splitter 252-D-04. The hydrocarbon stream from Unit 210 is first flashed in Drum 252-F-03. The absorber deethanizer has an upper and middle pump-around systems. The light gases are split into a fuel gas stream and a feed stream to CO₂ Removal Unit 213 in the SNG train. The deethanizer bottoms are sent to Debutanizer 252-D-03.

In 252-D-03, the propene/butene feed for Polymerization Unit 257 is obtained. The debutanizer uses a fired reboiler, 252-B-02.

The debutanizer bottoms, after cooling in Cooler 212-E-11, is charged to Splitter 252-D-04, in which naphtha is removed to obtain a heavier lean oil for recycling to Absorber Deethanizer 252-D-02. The tower bottoms (lean oil) and the overhead naphtha are recombined to become the feedstock for Hydrogenation Unit 253. The naphtha splitter also uses a fired reboiler, 252-B-03.

4.4.11 F.T. Product Hydrotreating Unit 253 (ZO-GEM-6979)

Purpose of the Unit is to saturate the olefins and to destroy the remaining alcohols and acids.

Technology Used

See Paragraph 3.4.17.

Process Description

The feed is pumped to reactor operating pressure (253-G-01), mixed with part of the hydrogen feed, heat exchanged against the reactor effluent in Exchanger 253-E-01, and finally heated to reactor operating temperature in Heater 253-B-01. Because of the high hydrogen consumption, Reactor 253-D-01 consists of multi-beds using a cold hydrogen quench for temperature control. A two-stage flash system is used: High Temp. Flash Drum 253-F-01 and Low Temp. Flash Drum 253-F-02. Liquid products are combined to become feed to Fractionation Unit 254. The hydrogen-rich gases from Drum 253-F-02 are split into hydrogen recycle and hydrogen feed for Poly Gasoline Hydrogenation Unit 259. The hydrogen recycle is compressed in Compressor 253-C-01, and then mixed with the make-up hydrogen. Reaction water collected in the low temperature drum is sent to Waste Water Treatment Unit 235.

Regeneration facilities for the hydrotreating catalyst are provided.

4.4.12 Hydrotreater Product Fractionation Unit 254 (ZO-GEM-6979)

Purpose of the Unit is to separate the hydrotreated product into a pentane/hexane stream for isomerization, a C₇/380 naphtha for reforming, a diesel oil product and a heavy fuel oil product.

Technology Used

A conventional 3-tower system is used.

Process Description

After heating in Heater 254-B-01, the hydrotreater product is charged to Main Fractionator 254-D-01, in which stripping steam is introduced at the bottom. The heavy fuel oil product is withdrawn from the bottom and after cooling sent to storage. A small amount of the heavy fuel oil, Stream 173.2, has been used in the final adjustment of the plant fuel balance. A side-cut is withdrawn to Steam Stripper 254-D-02 from which the diesel oil product, after cooling, is pumped to storage.

The light gases, butanes, pentanes/hexanes, naphtha and steam in the overhead product are cooled in Condensers 254-E-01 and 02, and separated in Reflux Drum 254-F-01. Light gases become a feed to Light Ends Recovery Unit 260. Water is pumped (254-G-09) to Waste Water Treatment Unit 235. The hydrocarbon liquid is split into reflux for the main fractionator and feed for Debutanizer 254-D-03.

In the debutanizer, essentially all the butanes are recovered and sent to Light Ends Unit 260. The debutanizer bottoms are charged to Naphtha Splitter 254-D-04. In the splitter, the pentanes and hexanes are fractionated from the heavy naphtha, or reformer charge and after cooling pumped (254-G-07) to Isomerization Unit 256. The cooled heavy naphtha is pumped (254-G-08) to Reforming Unit 255.

4.4.13 Catalytic Reforming Unit 255 (ZO-GEM-6980)

Purpose of the Unit is to increase the anti-knock quality of the C₇/380 naphtha.

Technology Used

Since its commercial introduction in 1950, catalytic reforming with platinum-based catalysts has become the major process for naphtha octane improvement (and aromatics production) in the petroleum industry. Reactions include: (1) dehydrogenation of naphthenes, (2) cyclization of straight-chain paraffins and subsequent dehydrogenation, (3) isomerization of paraffins and cyclopentane derivatives and (4) hydrocracking of paraffins and naphthenes. The formation of aromatics is the principal objective.

Catalytic reforming is carried out over a wide range of reaction conditions, the most important being pressure. Other variables include recycle gas rate, temperature, space velocity and catalyst particle size. Thermodynamically, dehydrogenation is favored by low pressure and high temperatures. These conditions, however, also favor the formation of coke. Fortunately, coke formation can be substantially suppressed by operating in an atmosphere of hydrogen without significant retardation of the dehydrogenation reactions. Typical reactor conditions are:

Pressure, psig	100-700
Recycle gas, SCF/Bbl	2,500-10,000
Temperature, °F	850-1,000
Space velocity, wt/hr/wt	1-3

In recent years, bimetallic catalysts - platinum plus other metal promoters, such as rhenium - have been commercialized. Their improved selectivity and activity permit reforming at lower pressures and higher severities, without shortening regeneration cycle times, than when using the previous standard platinum-on-alumina catalyst.

The principal problem in the design of catalytic reformers is heat balance. Reactions which produce aromatics are very endothermic. Although the exothermic hydrocracking reactions are partially offsetting, large amounts of heat must be supplied to the reaction zone. In fixed-bed units, the heat is supplied by employing multiple reactors in series with intermediate reheating of the reactants. As conditions dictate, a unit may be designed to operate either in a cyclic-or semi-regenerative mode of operation.

Most feedstocks are hydrotreated first to remove arsenic, sulfur and nitrogen compounds. Otherwise, these materials would poison the platinum catalyst. The cost of hydrotreating is easily justified by the extension of the reforming catalyst life.

Commercial processes are licensed by IFP, UOP Process, Houdry, Exxon, Chevron Research and Standard Oil (Indiana).

Process Description

The naphtha feed is pumped (255-G-04) to reactor operating pressure, mixed with the hydrogen-rich recycle gas, heat exchanged against Reactor 255-D-03 effluent in Exchanger 255-E-01 and heated to reactor operating temperature in Heater 255-B-01. The naphtha/hydrogen mixture then flows through three reactors in series, 255-D-01, 02 and 03, and is reheated before the second and third reactors in Heaters 255-B-02 and 03.

The third reactor effluent is cooled in 255-01, 02 and 03 and flashed in Drum 255-F-01. The hydrogen-rich overhead is split into a recycle gas stream and a feed stream to H₂ Purification Unit 261. The bottoms are pumped (255-G-03) to Stabilizer 255-D-04 for light ends fractionation. The butanes and lighter are sent to Light Ends Unit 260. After cooling, the stabilized reformat is sent to Gasoline Blending Unit 270.

Off-stream regeneration facilities are provided.

4.4.14 C₅/C₆ Isomerization Unit 256 (ZO-GEM-6981)

Purpose of the Unit is to increase the anti-knock quality of the pentanes and hexanes. Besides raising the pool octane, increasing the octane of the lighter gasoline constituents improves the octane balance between the front and back ends of the gasoline blend.

Technology Used

The first commercial isomerization units for the improvement of motor gasoline were announced in 1957. Isomerization requires the use of a catalyst. There is little heat evolved, and the reaction is limited by equilibrium and rate considerations. An economic balance is made between higher branched-chain ratios at low temperatures and higher reaction rates at high temperatures. Side reactions also must be considered in the selection of conditions.

Platinum-containing catalysts are favored in current commercial processes. They are promoted by the injection of trace amounts of organic chloride. The addition of a small amount of hydrogen suppresses the polymerization of the traces of olefins formed during the isomerization reaction and the formation of coke during hydrocracking reactions. Specifications for feedstock impurities, e.g., sulfur, water, benzene and C₇'s, vary considerably between the commercial processes. Licensors, however, have in recent years improved significantly impurity tolerances. Although operating conditions chosen generally allow a catalyst life of two or more years, should accidental deactivation occur, the catalyst can be easily regenerated and reactivated to full activity.

Typical reaction conditions are:

Temperature, °F	250-550
Pressure, psig	200-500
Space velocity, vol/hr/vol	1-5
H ₂ /hydrocarbon mole ratio	0.5-4

Processing conditions are generally milder for butane isomerization than for pentane or hexane isomerization.

Commercial processes are licensed by BP Trading, Shell Research and UOP Process.

Process Description

To maximize the octane improvement, Isomerization Unit 256 includes n-pentane recycling. The C_5/C_6 product from Unit 254 and the pentane products from isomerization are fed to Deisopentanizer 256-D-01. The overhead product, pentane isomerate, also contains the butanes and iso-pentane in the fresh feed and from 3 to 5 vol. % of n-pentane. After cooling, it is sent to Gasoline Blending Unit 270.

The n-pentane/hexane bottoms are mixed with recycle and make-up hydrogen, heated via Exchanger 256-D-04 and Heater 256-D-01 and passed over the catalyst in Reactor 256-D-02. The reactor effluent is cooled and flashed in Separator 256-F-02. A hydrogen-rich offgas is obtained for recycling via Compressor 256-G-01 and for hydrogen recovery in Unit 211.

The separator liquid is sent to Stabilizer 256-D-03 where the butanes and lighter are fractionated. The stabilizer offgases from Reflux Drum 256-F-03 after caustic washing are sent to H_2 Purification Unit 261, and the excess reflux to Light Ends Recovery Unit 260.

The stabilizer bottoms are charged to Depentanizer 256-D-04. The overhead product is pumped (256-G-04) to the deisopentanizer for recovery of the recycle n-pentane. The tower bottoms, hexane isomerate, after cooling, are sent to Gasoline Blending Unit 270.

4.4.15 Catalytic Polymerization Unit 257 (ZO-GEM-6982)

Purpose of the Unit is to polymerize the propene/butene mixture from the Fischer-Tropsch reaction into higher molecular weight compounds which are suitable for gasoline blending.

Technology Used

Catalytic polymerization has been a commercial process since about 1934 for the conversion of C_3 and C_4 olefins into a higher molecular weight poly gasoline. With a propene feed, trimers predominate and with a butene feed, dimers are the major products. Since the advent of catalytic reforming yielding a source of iso-butane, however, polymerization has generally been replaced by alkylation.

Olefin polymerization is catalyzed by acids and conversions are 85% or better. There is a large exothermic heat of reaction - about 400 Btu/lb of butene and about 670 Btu/lb of propene -, and a recycle stream is often used for temperature control. Commercial catalysts are mainly phosphoric acid on solids, such as kieselguhr and quartz chips.

The water content of the feed is critical. Optimum concentration, however, differs with the catalyst type, reactor operating conditions and feed composition. The olefin feed can be either "dried", saturated with water at a controlled temperature, or mixed with injected water.

The principal reactor design problem is temperature control. Multi-beds with liquid quench or tubular heat exchanger reactors generating steam are used. Typical operating conditions are:

Temperature, °F:	300 - 425
Pressure, psig:	400 - 1,500
Space velocity,	
Gal/hr/lb:	0.1 - 0.5

UOP Process and Chevron Research offer processes for licensing.

Process Description

The cat poly feed is pumped (257-G-01) to water saturation pressure and saturated in Water Settler 257-F-01. Because of the high concentration of olefins, only a portion of the fresh feed is mixed with the low olefin containing recycle stream, heated to reaction temperature in Exchanger 257-E-01 and Heater 257-E-02, and charged to the inlet of Reactor 257-D-01. The remaining cool fresh feed and, if needed, recycle are used as a quench for interbed cooling.

After passing through Acid K.O. Drum 257-F-02 and Heat Exchanger 257-E-01, the reactor effluent is split into the recycle stream and the feed for Debutanizer 257-D-02. The overhead product, consisting of unreacted olefins, iso-butane and inert propane and n-butane, is cooled and pumped to Alkylation Unit 258. The poly gasoline (bottoms) is sent to Splitter 257-D-03, where it is separated at the C₉ - C₁₀ boiling range. The light poly gasoline is pumped (257-G-03) to Gasoline Blending Unit 270 and the heavy poly gasoline is pumped (257-G-04) to Hydrogenation Unit 259.

4.4.16 H.F. Alkylation Unit 258 (ZO-GEM-6982)

Purpose of the Unit is to increase the gasoline yield by the catalytic alkylation of iso-butane and unpolymerized C₃ and C₄ olefins.

Technology Used

See Paragraph 3.4.19.

Process Description

Except for an additional iso-butane-rich feed stream, identical to H.F. Alkylation Unit 153. See Paragraph 3.4.19.

4.4.17 Poly Gasoline Hydrogenation Unit 259 (ZO-GEM-6983)

Purpose of the Unit is to saturate the olefins in the heavy cat poly gasoline.

The olefin target concentration for the 10 RVP gasoline product is 20 vol. % maximum. To meet this specification, the poly gasoline is split and the heavy gasoline is hydrogenated. Preliminary gasoline blending indicated that the C₉ and higher olefins should be hydrogenated and the unit has been sized accordingly. Final blending results, however, show that C₉ olefin hydrogenation is not necessary. Consequently, Unit 259 is sized conservatively.

Technology Used

See Paragraph 3.4.17.

Process Description

The heavy cat poly gasoline after mixing with the make-up/recycle hydrogen is heated via the reactor effluent in Exchanger 259-E-01 and brought-up to the operating temperature by Heater 259-B-01. Reactor 259-D-01 uses multi-beds with hydrogen quench for temperature control. The effluent is cooled in Coolers 259-E-02 and 03, and flashed in Drum 259-F-01. The overhead is split into recycle and offgas streams.

The bottoms liquid flows to Stabilizer 259-D-02. After cooling the stabilizer offgases are combined with the offgases from Flash Drum 259-F-01 and sent to H₂ Purification Unit 261. The cooled stabilized heavy gasoline is sent to Gasoline Blending Unit 270.

Catalyst regeneration facilities are provided.

4.4.18 Light Ends Recovery Unit 260 (ZO-GEM-6980)

Purpose of the Unit is to separate the light gasoline and iso-butane from the hydrocarbon-rich offgases.

Process Description

Offgas streams from Units 254, 255 and 256 are first flashed in Drum 260-F-02. The overhead vapors are compressed (260-C-01) and the liquid is pumped (260-G-01) to the operating pressure of Deethanizer 260-D-01.

The low-pressure offgas from Unit 254 is first compressed and then flashed with the offgas from Unit 261 in Drum 260-F-01. The drum liquid is charged to Deethanizer 260-D-01.

The deethanizer overhead is combined with Drum 260-F-01 overhead and Stabilizer 255-D-04 uncondensed gases and sent to the fuel gas header. The feed to Debutanizer 260-D-02 is the deethanizer bottoms. In the debutanizer, the additional iso-butane feed for Unit 258 is fractionated from the light gasoline which is sent to Gasoline Blending Unit 270.

4.4.19 H₂ Purification Unit 261 (ZO-GEM-6984)

Purpose of the Unit is to obtain from the hydrogen-rich offgases a hydrogen make-up stream of 90% purity.

Technology Used

Commercial cryogenic technology, such as offered by Linde Div. of Union Carbide Corp., is used.

Process Description

Two of the three feed streams require compressing (261-C-01 and 02) before charging to the system. The combined feed is first cooled in Cooler 261-E-01 by the +32°F refrigerant from Refrigeration Unit 241 and flashed in Drum 261-F-01 for heavy hydrocarbon and water removal. The gases then pass through the Molsieve Section 261-D-01 for final drying and CO₂ removal. In Cold Box Section 261-D-02, the make-up hydrogen is recovered and sent to Unit 211 for final compression. The cold box hydrocarbon gases are compressed in Compressor 261-C-03 and mixed with the liquids separated in Flash Drum 261-F-01. The resultant mixture is a feed for Light Ends Recovery Unit 260.

4.4.20 Alcohol Recovery Unit 262 (ZO-GEM-6985)

Purpose of the Unit is to prepare a marketable alcohol mixture product and to recover methanol for make-up to Rectisol Unit 205.

Technology Used

Because of the limited alcohol information available to MRDC, the design of this unit is general in nature. Distillation is used to separate the bulk of the water and acids from the raw alcohols. Methanol is then fractionated from the alcohol mixture. A benzene azeotropic system is used for final water removal. The final step is the hydrogenation of the aldehydes and ketones to improve the marketing value of the alcohol mixture.

Process Description

The alcohol/acid water mixture from F-T Synthesis Unit 250 flows to Fractionator 262-D-01 in which the alcohols are distilled from the acidic water. Because alcohol-water azeotropic mixtures occur, a portion of the water distills overhead with the alcohols. The alcohols, together with a heavy alcohol recycle stream from Methanol Tower 262-D-03, are next charged to Alcohol Splitter Tower 262-D-02 from which a crude methanol overhead product is obtained. The crude methanol is redistilled in the methanol tower to obtain an acceptable make-up material for Gas Purification Unit 205. The higher boiling alcohols (bottoms) are returned to the alcohol splitter tower.

The bottoms stream from Alcohol Splitter Tower 262-D-02 is fed to Heavy Alcohol Tower 262-D-04 into which benzene is added to remove, via the benzene-water azeotrope, the remaining water. The azeotropic overhead is pumped to Tower 262-D-05, where 50 psig steam is used to strip the benzene which is then returned to Tower 262-D-04.

The alcohol bottoms stream from Alcohol Tower 262-D-04 is pumped to Hydrogenation Section 262-D-06 where the aldehyde and ketone impurities are hydrogenated to their corresponding alcohols. Make-up hydrogen is supplied from H₂ Recovery Unit 211 and the purge gas is returned to the same unit. A small portion of the upgraded alcohol mixture is sent to Hydrocarbon Recovery Unit 210 for the prevention of ice formation during the cold hydrocarbon washing operation.

The waste water streams from Towers 262-D-01 and 05 are pumped (262-G-02 and 09) after cooling (262-E-13) to Waste Water Treatment Unit 235.

4.4.21 Unit Capacity Summary

The capacities of the principal process units are summarized below in conventional flow rates:

<u>Unit No.</u>	<u>Name</u>	<u>--Capacity, Unit/SD--</u>	
		<u>Input</u>	<u>Output</u>
201	Gasification	22.8 MT coal	1,755 MMSCF
202	Raw Gas Shift	331 MMSCF	-
205	Gas Purification	1,062 MMSCF	742 MMSCF
210	Hydrocarbon Recovery	277 MMSCF	235 MMSCF
250	F-T Synthesis	742 MMSCF	18.1 MBbl
251	Naphtha Hydro-treating	1.3 MBbl	-
253	F-T Prod. Hydro-treating	11.4 MBbl	-
255	Catalytic Reforming	4.3 MBbl	-
256	C ₅ /C ₆ Isomerization	4.0 MBbl	-
257	Polymerization	6.5 MBbl	-
258	Alkylation	2.4 MBbl (550 Bbl/SD alkylate)	-
259	Poly Gaso. Hydro-generation	1.2 MBbl	-
260	Light Ends Recovery	1.0 MBbl	-
261	H ₂ Purification	8 MMSCF	6 MMSCF
262	Alcohol Recovery	510 Mlb	-
	SNG Train	235 MMSCF	173 MMSCF

TYPE NO.	NAME	QTY	APPROX. - PAGE 1-2, LENGTH 2-4
210-C-01	OFF-GAS COMPRESSOR	1	0
210-D-01	1.000000	1	0
210-D-02	1.000000	1	0
210-E-01	HEAT EXCHANGER	1	0
210-E-02	HEAT EXCHANGER	1	0
210-E-03	HEAT EXCHANGER	1	0
210-E-04	HEAT EXCHANGER	1	0
210-E-05	HEAT EXCHANGER	1	0
210-E-06	HEAT EXCHANGER	1	0
210-E-07	HEAT EXCHANGER	1	0
210-E-08	HEAT EXCHANGER	1	0
210-E-09	HEAT EXCHANGER	1	0
210-E-10	HEAT EXCHANGER	1	0
210-F-01	REGULATOR	1	0
210-F-02	REGULATOR	1	0
210-G-01	REGULATOR	1	0
210-G-02	REGULATOR	1	0
210-G-03	REGULATOR	1	0
210-H-01	REGULATOR	1	0
210-H-02	REGULATOR	1	0
210-H-03	REGULATOR	1	0
210-H-04	REGULATOR	1	0
210-H-05	REGULATOR	1	0
210-H-06	REGULATOR	1	0
210-H-07	REGULATOR	1	0
210-H-08	REGULATOR	1	0
210-H-09	REGULATOR	1	0
210-H-10	REGULATOR	1	0

NOTES:
 STEAMS WITHOUT STREAM NUMBERS
 ARE NOT INCLUDED IN THE OVERALL
 MATERIAL BALANCE.

Project No.	13.777 VI	Client	GEM	Unit	210
Drawn By		Checked By		Approved By	
PROCESS FLOW DIAGRAM			Z0-GEM-6973		
HYDROCARBON RECOVERY			HEPTANE WASH		
UNIT 210			BASE CASE II

ITEM NO.	NAME	QTY	APPROX. COST (\$)
211-C-01	PRESSURE SWING OFFGAS RECOMPRESSOR	1	0
211-C-02	UNIT 211 HYDROGEN COMPRESSOR	1	0
211-C-03	UNIT 262 HYDROGEN COMPRESSOR	1	0
211-C-04	UNIT 281 HYDROGEN COMPRESSOR	1	0
211-D-01	PRESSURE SWING OFFGAS HOLDER	4	0
211-F-01	PRESSURE SWING OFFGAS HOLDER	1	0
251-B-01	NAPHTHA REACTOR	1	0
251-C-01	HYDROGEN RECYCLE COMPRESSOR	1	1
251-D-01	NAPHTHA HYDROTREATING REACTOR	1	0
251-D-02	AMINE WASH TOWER	1	0
251-D-03	NAPHTHA STABILIZER	1	0
251-E-01	REACTOR EFFLUENT/REACTOR FEED H.E.	1	0
251-E-02	REACTOR EFFLUENT COOLER	1	0
251-E-03	REACTOR EFFLUENT TRIM COOLER	1	0
251-E-04	STABILIZER REBOILER	1	0
251-E-05	STABILIZER OVER CONDENSER	1	0
251-E-06	STABILIZER OVER TRIM CONDENSER	1	0
251-E-07	STABILIZER NAPHTHA COOLER	1	0
251-E-08	STABILIZER NAPHTHA TRIM COOLER	1	0
251-F-01	NAPHTHA FLASH DRUM	1	0
251-F-02	AMINE WASH OVER H.E. DRUM	1	0
251-F-03	STABILIZER REFLEX DRUM	1	0
251-G-01	NAPHTHA FEED PUMP	1	1
251-G-02	STABILIZER REBOILER PUMP	1	1
251-G-03	STABILIZER REFLEX PUMP	1	1
251-G-04	AMINE CIRCULAT PUMP	1	1
251-G-05	NAPHTHA HYDROTREATING WATER PUMP	1	1

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE.

GEM		Lurgi	
00-0865		Lurgi Process Technology GmbH	
PROCESS FLOW DIAGRAM		Z0-GEM-6974	
H ₂ RECOVERY (PRESSURE SWING)		HYDROTREATING STUDY	
& NAPHTHA HYDROTREATING		EF-76-C-01-2447	
UNITS 211/251		BASE CASE II

ITEM NO.	NAME	OPER. APPROX. REPAIR I.D. POWER S-Y
212-B-01	REDUCTION SYSTEM HEATER	1 + 0 NOT ON PFD
212-C-01	RECYCLE COMPRESSOR	2 + 0
212-D-01	REACTOR	2 + 0 11'-4" / 8'-8"
212-D-02	CLEAN-UP REACTOR	2 + 0 8'-0" / 10'-8"
212-E-01	STEAM GENERATOR	2 + 0
212-E-02	RECYCLE/BFW H.E.	2 + 0
212-E-03	FIRST FEED/PRODUCT H.E.	1 + 0
212-E-04	SECOND FEED/PRODUCT H.E.	1 + 0
212-E-05	THIRD FEED/PRODUCT H.E.	1 + 0
212-E-06	PRODUCT COOLER	1 + 0
212-E-07	PRODUCT TRIM COOLER	1 + 0
212-E-08	REDUCTION SYSTEM COOLER	1 + 0 NOT ON PFD
212-F-01	FIRST E.O. DRUM	1 + 0
212-F-02	SECOND E.O. DRUM	1 + 0
212-F-03	THIRD E.O. DRUM	1 + 0
212-F-04	REDUCTION SYSTEM E.O. DRUM	1 + 0 NOT ON PFD

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE.

Project No.	12777 VI	GEM	LURGI
Drawn by		00-0865	Lurgi Mineralabs GmbH
Checked by			Frankfurt (Main)
Scale	1:1		
Sheet No.	1		
Process	PROCESS FLOW DIAGRAM		
Unit	UNIT 212		
Case	BASE CASE II		
Study	METHANATION		
Ref.	20-GEM-6975		
Rev.	REV. 01-2447

EQUIPMENT LIST		
ITEM NO.	NAME	SPR. OFFICE. -S&P-1
213-D-01	CO ₂ ABSORBER	1-0 84-17/91
213-D-02	REGENERATOR	1-0 84-17/91
213-E-01	LEAF/RICH S.S. COOLER	1-0
213-E-02	TRIM COOLER	1-0
213-E-03	REGENERATOR OF NO CONDENSER	1-0
213-E-04	REGENERATOR OF NO CONDENSER	1-0
213-E-05	REGENERATOR OF NO CONDENSER	1-0
213-E-06	REGENERATOR OF NO CONDENSER	1-0
213-E-07	CO ₂ ABSORBER WASH WATER COOLER	1-0
213-F-01	FILTER	1-0
213-F-02	REGENERATOR REFLEX DRUM	1-0
213-F-03	REGENERATOR OFFGAS D.O. DRUM	1-0
213-G-01	DILUTION PUMP	1-1
213-G-02	REGENERATOR REFLEX PUMP	1-1
214-C-01	SNG COMPRESSOR	1-0
214-D-01	WATER ABSORBER	1-0
214-D-02	TRIM SUBGENERATOR	1-0
214-E-01	TRIM COOLER	1-0
214-E-02	SNG COMPRESSOR COOLER	1-0
214-F-01	TRIM DRUM	1-0
214-F-02	SNG COMPRESSOR SUCTION DRUM	1-0
214-G-01	TRIM PUMP	1-1

Project No.	27777 VI	GEM	00-0865	LCRGE
Drawn By				
Checked By				
Approved By				
Process Flow Diagram	CO ₂ REMOVAL & SNG DRYING & COMPR.			20-SEM-6976
Unit	UNIT 213/214			20-SEM-6976
Base Case	BASE CASE II			20-SEM-6976

EQUIPMENT LIST			
ITEM NO.	NAME	OPER. #	APPROX. I.D. LENGTH Y.F.
250-C-01	SYNGAS & RECYCLE COMPRESSOR	2-0	
250-D-01	F-T REACTOR SYSTEM	4-1	10'-6" 100'
250-D-02	F-T EFFLUENT WASH TOWER	4-0	18'-6" 50'
250-D-03	F-T VAPOR COLD WASH TOWER	2-0	20'-6" 50'
250-D-04	LIGHT OIL WASH TOWER	1-0	13'-6" 30'
250-E-01	F-T FLEED/EFFLUENT HEAT EXCHANGER	1-0	
250-E-02	F-T WASTE HEAT EXCHANGER	1-0	
250-E-03	200 PSIG STEAM GENERATOR	1-0	
250-E-04	F-T WASTE HEAT CONDENSER	1-0	
250-E-05	F-T WASTE HEAT TRIP CONDENSER	1-0	
250-E-06	F-T WASTE HEAT TRIP CONDENSER	1-0	
250-E-07	50 PSIG STEAM GENERATOR	1-0	
250-E-08	F-T WASTE HEAT EXCHANGER	1-0	
250-F-01	COMPRESSOR SECTIONAL DRAIN	2-0	
250-F-02	REF. OIL SHOCK DRAIN	1-0	
250-G-01	REF. OIL CIRCULATION PUMP	2-1	
250-G-02	F-T EFFLUENT SIDE STREAM PUMP	4-2	
250-G-03	F-T EFFLUENT WASH TOWER	2-1	
250-G-04	WASH WATER PUMP	1-1

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE.

Project No.	14.777	GEM	00-0865	LURG
Drawn By				
Checked By				
Approved By				
Project Name	PROCESS FLOW DIAGRAM			20-GEM-6977
Project No.	F-T SYNTHESIS			DC/EMCA STUDY
Project No.	UNIT 250			CF-76-C-01-2447
Project No.	BASE CASE II

EQUIPMENT LIST				
ITEM NO.	NAME	OPER. SPACE	APPROX. I.D., LENGTH	FEET
252-B-01	HEAVY OIL REFLUX REBOILER	1 x 0		
252-B-02	DEBUTANIZER REBOILER	1 x 0		
252-B-03	SPLITTER REBOILER	1 x 0		
252-C-01	VACUUM TOWER OVER BLOWER	1 x 0		
252-D-01	HEAVY OIL VACUUM TOWER	1 x 0	4'-0"	80'
252-D-02	ABSORBER/DEBUTANIZER (AB/DE)	1 x 0	7'-0"	120'
252-D-03	DEBUTANIZER	1 x 0	7'-0"	117'
252-D-04	KAPHTHA SPLITTER	1 x 0	11'-0"	41'
252-E-01	VACUUM TOWER P/A COOLER	1 x 0		
252-E-02	VACUUM TOWER REFLUX COOLER	1 x 0		
252-E-03	VACUUM TOWER OVER CONDENSER	1 x 0		
252-E-04	LEAK OIL TRIM COOLER	1 x 0		
252-E-05	AB/DE FLD. LEAK OIL H.E.	1 x 0		
252-E-06	AB/DE TOP P/A COOLER	1 x 0		
252-E-07	AB/DE BOTTOM P/A COOLER	1 x 0		
252-E-08	AB/DE REBOILER	1 x 0		
252-E-09	DEBUTANIZER OVER CONDENSER	1 x 0		
252-E-10	DEBUTANIZER OVER TRIM CONDENSER	1 x 0		
252-E-11	SPLITTER FLD. COOLER	1 x 0		
252-E-12	SPLITTER OVER CONDENSER	1 x 0		
252-E-13	LEAK OIL COOLER	1 x 0		
252-E-14	SPLITTER BOTTOMS COOLER	1 x 0		
252-F-01	HEAVY OIL FLASH DRUM	1 x 0		
252-F-02	VACUUM TOWER OVER ACCUMULATOR	1 x 0		
252-F-03	REFLUX BASE OFF STREAM	1 x 0		
252-F-04	FLASH DRUM	1 x 0		
252-F-05	DEBUTANIZER REFLUX DRUM	1 x 0		
252-F-06	SPLITTER REFLUX DRUM	1 x 0		
252-G-01	VACUUM TOWER P/A PUMP	1 x 1		
252-G-02	VACUUM TOWER REFLUX PUMP	1 x 1		
252-G-03	LEAK OIL FEED PUMP	1 x 1		
252-G-04	AB/DE TOP P/A PUMP	1 x 1		
252-G-05	AB/DE BOTTOM P/A PUMP	1 x 1		
252-G-06	AB/DE REBOILER PUMP	1 x 1		
252-G-07	DEBUTANIZER REFLUX PUMP	1 x 1		
252-G-08	DEBUTANIZER REBOILER PUMP	1 x 1		
252-G-09	SPLITTER REFLUX PRODUCT PUMP	1 x 1		
252-G-10	SPLITTER REBOILER PUMP	1 x 1		
252-G-11	SPLITTER BOTTOMS PUMP	1 x 1		
252-G-12	LEAK OIL PUMP	1 x 1		
252-G-13	VACUUM TOWER ACCUM. PUMP	1 x 1		
252-G-14				
252-G-15

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE.

Project No.	15777 VI	GEM	Lurgi
Client		00-0865	Large Chemicals Division
Design No.			
Drawn By			
Checked By			
Approved By			
Process Flow Diagram	F-T PRODUCT FRACTIONATION		
Unit 252	BASE CASE II		
Design No.	70-GEM-6978		
MECH. ERDA STUDY	EF-76-C-01-2447		
Supervised By			
Engineer's Code

EQUIPMENT LIST				
ITEM NO.	NAME	OPER.	APPROX. I.D. (INCHES)	TYPE
253-B-01	HYDROTREATER FEED HEATER	1 + 0		
253-C-01	HYDROTREATER RECYCLE COMPRESSOR	1 + 0		
253-D-01	HYDROTREATER REACTOR	1 + 0	6'-0"	30'
253-E-01	HYDROTREATER FEED/EFFLUENT H.E.	1 + 0		
253-E-02	HT FLASH OVD/LT FLASH HTGR. H.E.	1 + 0		
253-E-03	HT FLASH OVD CONDENSER	1 + 0		
253-E-04	HT FLASH OVD TRIM CONDENSER	1 + 0		
253-E-05	QUENCH HYDROGEN COOLER	1 + 0		
253-F-01	HIGH TEMP FLASH DRUM	1 + 0		
253-F-02	LOW TEMP FLASH DRUM	1 + 0		
253-F-03	HYDROTREATER FEED PUMP	1 + 0		
254-B-01	MAIN FRACTIONATOR FEED HEATER	1 + 0		
254-D-01	MAIN FRACTIONATOR	1 + 0	8'-0"	80'
254-D-02	DISTILLATE FUEL OIL STRIPPER	1 + 0	6'-0"	60'
254-D-03	DEBUTANIZER	1 + 0	6'-0"	71'
254-D-04	RAFFINATE SPLITTER	1 + 0	6'-0"	30'
254-E-01	MAIN FRACTIONATOR FEED CONDENSER	1 + 0		
254-E-02	MAIN FRACTIONATOR OVD TRIM CONDENSER	1 + 0		
254-E-03	L/O PRODUCT/DEBUTANIZER FEED H.E.	1 + 0		
254-E-04	DPO PRODUCT COOLER	1 + 0		
254-E-05	EPO PRODUCT/DEBUTANIZER FEED H.E.	1 + 0		
254-E-06	EPO PRODUCT COOLER	1 + 0		
254-E-07	DEBUTANIZER FEED HEATER	1 + 0		
254-E-08	DEBUTANIZER OVD CONDENSER	1 + 0		
254-E-09	DEBUTANIZER OVD TRIM CONDENSER	1 + 0		
254-E-10	DEBUTANIZER REBOILER	1 + 0		
254-E-11	KAPHTMA SPLITTER OVD CONDENSER	1 + 0		
254-E-12	KAPHTMA SPLITTER REBOILER	1 + 0		
254-E-13	PETRAK/REKANS TRIM COOLER	1 + 0		
254-E-14	KAPHTMA SPLITTER HTS COOLER	1 + 0		
254-F-01	MAIN FRACTIONATOR REFLUX DRUM	1 + 0		
254-F-02	DEBUTANIZER REFLUX DRUM	1 + 0		
254-F-03	KAPHTMA SPLITTER REFLUX DRUM	1 + 0		
254-G-01	MAIN FRACTIONATOR REFLUX PUMP	1 + 1		
254-G-02	DEBUTANIZER FEED PUMP	1 + 1		
254-G-03	DPO PRODUCT PUMP	1 + 1		
254-G-04	EPO PRODUCT PUMP	1 + 1		
254-G-05	DEBUTANIZER REFLUX PUMP	1 + 1		
254-G-06	KAPHTMA SPLITTER REFLUX PUMP	1 + 1		
254-G-07	KAPHTMA SPLITTER OVD PRODUCT PUMP	1 + 1		
254-G-08	KAPHTMA SPLITTER HTS PUMP	1 + 1		
254-G-09	WASTE WATER PUMP	1 + 1

NOTES:
 STREAMS WITHOUT STREAM NUMBERS
 ARE NOT INCLUDED IN THE OVERALL
 MATERIAL BALANCE.

Project No.	15777-VI	GEM	LURGI
Client		00-0865	Energy Management GmbH
Drawn by			Frankfurt (Main)
Check by			
Approved by			
Scale			
Sheet No.			
Process Name	PROCESS FLOW DIAGRAM		
Unit Name	F-T PROD. HYDROTREATER & FRACTIONATION		
Base Case	UNITS 253/254 BASE CASE II		
Study No.	20-GEM-6979		
Study Title	F-T PROD. HYDROTREATER & FRACTIONATION		
Study Date	1976-01-24/27		
Study By	Lurgi AG		
Study For	Lurgi AG

ITEM NO.	NAME	QTY	APPROX.	UNIT
256-B-01	ISOMERIZATION REACTOR REATER	1	0	
256-C-01	ISOMERIZATION RECYCLE COMPRESSOR	1	0	
256-D-01	DEISOPENTANIZER	1	0	178'
256-D-02	ISOMERIZATION REACTOR	1	0	87'
256-D-03	STABILIZER	1	0	180'
256-D-04	DEPENTANIZER	1	0	120'
256-E-01	DEISOPENTANIZER OVER CONDENSER	1	0	
256-E-02	DEISOPENTANIZER OVER TRIM CONDENSER	1	0	
256-E-03	DEISOPENTANIZER REBOILER	1	0	
256-E-04	ISOMERIZATION REACTOR FEED/	1	0	
256-E-05	ISOMERIZATION REACTOR EFFLUENT	1	0	
256-E-06	COOLER	1	0	
256-E-07	ISOMERIZATION REACTOR EFFLUENT	1	0	
256-E-08	TRIM COOLER	1	0	
256-E-09	STABILIZER FEED/BOTTOMS S.E.	1	0	
256-E-10	STABILIZER OVER CONDENSER	1	0	
256-E-11	STABILIZER OVER TRIM CONDENSER	1	0	
256-E-12	STABILIZER REBOILER	1	0	
256-E-13	DEPENTANIZER OVER CONDENSER	1	0	
256-E-14	DEPENTANIZER REBOILER	1	0	
256-E-15	DEPENTANIZER BOTTOMS COOLER	1	0	
256-E-16	DEPENTANIZER BOTTOMS TRIM COOLER	1	0	
256-F-01	DEISOPENTANIZER REFLUX/PRODUCT	1	0	
256-F-02	ISOMERIZATION REFLUX SEPARATOR	1	0	
256-F-03	STABILIZER REFLUX DRUM	1	0	
256-F-04	DEPENTANIZER REFLUX DRUM	1	0	
256-G-01	DEISOPENTANIZER REFLUX/PRODUCT	1	0	
256-G-02	STABILIZER REFLUX PUMP	1	0	
256-G-03	DEPENTANIZER REFLUX PUMP	1	0	
256-G-04	REACTOR RECYCLE PUMP	1	0	
256-G-05	REACTOR FEED PUMP	1	0	
256-G-06	HEXANE, ISOMERATE PUMP	1	0	

◊ STREAM NUMBER
 ○ PRESSURE
 □ TEMPERATURE
 ▭ FLOW RATE OFF GAS
 ▭ FLOW RATE STEAM AND LIQUIDS

NOTES:
 STREAMS WITHOUT STREAM NUMBERS
 ARE NOT INCLUDED IN THE OVERALL
 MATERIAL BALANCE

217.77	VI	GEM	00-0865	LURGI
PROCESS FLOW DIAGRAM		ZD-GEM-6981		
C5/C6 ISOMERIZATION		M-DC/EPDA STUDY		
UNIT 256		EF-76-C-01-2447		
BASE CASE II

ITEM NO.	NAME	EQUIPMENT DATA		
		OPER. CAPACITY	APPROX. I.D.	LENGTH
259-B-01	FEED REACTOR	1 + 0		
259-C-01	RECYCLE COMPRESSOR	1 + 1		
259-D-01	HYDROGENATION REACTOR	1 + 0	2'-6"	30'
259-D-02	STABILIZER	1 + 0	2'-6"	70'
259-E-01	HYDROGENATION FEED/EFFLUENT E.C.	1 + 0		
259-E-02	HYDROGENATION EFFLUENT COOLER	1 + 0		
259-E-03	HYDROGENATION EFFLUENT THERMOCOOLER	1 + 0		
259-E-04	STABILIZER OVERHEAD CONDENSER	1 + 0		
259-E-05	STABILIZER OVERHEAD TANK CONDENSER	1 + 0		
259-E-06	STABILIZER REBOILER	1 + 0		
259-E-07	STABILIZER BOTTOMS COOLER	1 + 0		
259-E-08	STABILIZER BOTTOMS THERMOCOOLER	1 + 0		
259-F-01	HYDROGENATION EFFLUENT PUMP	1 + 0		
259-F-02	STABILIZER REFLUX PUMP	1 + 0		
259-G-01	FEED PUMP	1 + 1		
259-G-02	STABILIZER REFLUX PUMP	1 + 1		
259-G-03	STABILIZER REBOILER PUMP	1 + 1

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE

Drawn by	Checked by	Reviewed by	Approved by
12777 VI			
Project	GEM	00-0865	LURGI
Client	Lurgi AG, Ludwigshafen (Germany)		
Drawn by	Z0-GEM-6983		
Project	POLY GASOLINE HYDROGENATION		
Client	EF-76-C-01-2447		
Drawn by	AD		
Project	UNIT 259		
Client	BASE CASE II

EQUIPMENT LIST

ITEM NO.	NAME	OPER.	APPROX.
261-C-01	COLD BOX MAIN FEED COMPRESSOR	1 + 1	
261-C-02	COLD BOX SEC. FEED COMPRESSOR	1 + 1	
261-C-03	COLD BOX MC PRODUCT COMPRESSOR	1 + 1	
261-D-01	MOL SIEVE ADSORBER	2 + 0	
261-D-02	COLD BOX SECTION	1 + 0	
261-E-01	MOL SIEVE FEED CHILLER	1 + 0	
261-E-02	COLD BOX MC PRODUCT COOLER	1 + 0	
261-F-01	MOL SIEVE FLASH DRUM	1 + 0	
261-F-02	COLD BOX MC PRODUCT E.O. DRUM	1 + 0

NOTES:
STREAMS WITHOUT STREAM NUMBERS
ARE NOT INCLUDED IN THE OVERALL
MATERIAL BALANCE.

Project No.	12777	Rev.	VI	GEM	00-0865	LURGI
Client	Lurgi Mineralprodukte GmbH					
Project Name	Z0-GEM-F984					
Process	H2 PURIFICATION					
Unit	UNIT 261					
Case	BASE CASE II

EQUIPMENT LIST				
Item No.	Name	Oper. Spare	Approx. I.D.	Length T.
262-D-01	ACID WATER FRACTIONATOR	1 + 0	12'-0"	140'
262-D-02	ALCOHOL SPLITTER	1 + 0	2'-0"	80'
262-D-03	METHANOL TOWER	1 + 0	2'-0"	70'
262-D-04	MTY ALCOHOL TOWER	1 + 0	18'-0"	70'
262-D-05	ALCOHOL RECOVERY STEAM STRIPPER	1 + 0	8'-0"	70'
262-E-01	ACID WATER FRACTIONATOR REBOILER	1 + 0		
262-E-02	ACID WATER FRACTIONATOR CONDENSER	1 + 0		
262-E-03	ACID WATER FRACTIONATOR OVER-TRIM CONDENSER	1 + 0		
262-E-04	ALCOHOL SPLITTER REBOILER	1 + 0		
262-E-05	ALCOHOL SPLITTER OVER-TRIM CONDENSER	1 + 0		
262-E-06	METHANOL TOWER REBOILER	1 + 0		
262-E-07	METHANOL TOWER OVER-TRIM CONDENSER	1 + 0		
262-E-08	METHANOL TOWER COOLER	1 + 0		
262-E-09	MTY ALCOHOL TOWER REBOILER	1 + 0		
262-E-10	MTY ALCOHOL TOWER OVER-TRIM CONDENSER	1 + 0		
262-E-11	MTY ALCOHOL TOWER STEAM COOLER	1 + 0		
262-E-12	ACID WATER COOLER	1 + 0		
262-E-13	ACID WATER COOLER	1 + 0		
262-F-01	ACID WATER TOWER REFLUX/PRODUCT PUMP	1 + 1		
262-F-02	ALCOHOL SPLITTER REFLUX/PRODUCT PUMP	1 + 1		
262-F-03	ALCOHOL SPLITTER REFLUX/PRODUCT PUMP	1 + 1		
262-F-04	MTY ALCOHOL TOWER REFLUX/PRODUCT PUMP	1 + 1		
262-F-05	MTY ALCOHOL TOWER REFLUX/PRODUCT PUMP	1 + 1		
262-F-06	MTY ALCOHOL TOWER REFLUX/PRODUCT PUMP	1 + 1		
262-F-07	STEAM STRIPPER REFLUX/PRODUCT PUMP	1 + 1		
262-F-08	ALCOHOL HYDROLYSIS REFLUX/PRODUCT PUMP	1 + 1		
262-F-09	STEAM CONDENSATE PUMP	1 + 1		

- ◇ STREAM NUMBER
- PRESSURE
- TEMPERATURE
- ▭ FLOW RATE MTY GAS
- ▭ FLOW RATE STEAM AND LIQUIDS

NOTES:
STREAMS WITHOUT STREAM NUMBERS ARE NOT INCLUDED IN THE OVERALL MATERIAL BALANCE.

Project No.	22777	Client	GEM	Project No.	00-0865
Design	22777	Client	Lurgi	Project No.	00-0865
Design	22777	Client	Lurgi	Project No.	00-0865
Design	22777	Client	Lurgi	Project No.	00-0865
Design	22777	Client	Lurgi	Project No.	00-0865
Design	22777	Client	Lurgi	Project No.	00-0865
Design	22777	Client	Lurgi	Project No.	00-0865
Design	22777	Client	Lurgi	Project No.	00-0865
Design	22777	Client	Lurgi	Project No.	00-0865
Design	22777	Client	Lurgi	Project No.	00-0865

PROCESS FLOW DIAGRAM
ALCOHOL RECOVERY

UNIT 262 BASE CASE II

BASE CASE II
4.5 OFFSITE UNITS

Majority of the offsite units in Base Case II are very similar to those in Base Case I. Consequently, descriptions are not repeated; differences, principally capacities, are discussed below.

Unless stated otherwise, units are shown on Block Flow Diagram ZO-GEM-6988 in Sub-Section 4.4. In addition, units such as boilers, generators and water treaters are also included in Steam/BFW Balance ZO-GEM-6986 located in Sub-Section 4.6.

4.5.1 Oxygen Production Unit 221

Identical to Unit 121.

4.5.2 Boiler Unit 222

Differences are as follows:

Main Fuel

Fine coal quantity (composition listed in Sub-Section 2.3)	414.7 Mlb/hr
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Supplementary Fuels

- Oil quantity (from Gas Liquor Separation Unit 207)	46.49 Mlb/hr
- Vacuum tower residue (from Fractionation Unit 252)	
Quantity	2.0 Mlb/hr
Low heating value	17,750 Btu/lb

Incineration Fuel

Sour offgases (mainly from Sulfur Recovery Unit 206)	
Quantity	1,827.7 Mlb/hr
Low heating value	243 Btu/lb

NOTE: The heat content of these gases (30 Btu/SCF) has not been included in the boiler thermal balance.

<u>Normal Capacity</u>	3,111.3 Mlb/hr
------------------------	----------------

4.5.3 Main Superheater Unit 223

Capacity is changed as follows:

<u>Normal Capacity</u>	3,111.3 Mlb/hr
------------------------	----------------

4.5.4 Superheater Unit 224

Duties are as follows:

	<u>200 psig Steam</u>	<u>100 psig Steam</u>
Inlet pressure, psig	200	100
Outlet pressure, psig	185	85

	<u>200 psig Steam</u>	<u>100 psig Steam</u>
Inlet temperature, °F	387	338
Outlet temperature, °F	586	469

Fuel consists of a portion of the oil recovered from Gas Liquor Separator Unit 107:

Oil quantity 5.22 Mlb/hr

Normal Capacity

Coil 1: 200 psig steam superheating	78.7 Mlb/hr
Coil 2: 100 psig steam superheating	777.1 Mlb/hr

4.5.5 Electrostatic Stack Gas Precipitator Unit 225

Normal Capacity 216.9 Mlb-mol/hr

4.5.6 Stack Gas Clean-Up Unit 226

Normal Capacity 216.9 Mlb-mol/hr containing
48.1 lb-mol/hr SO₂

4.5.7 Instrument and Plant Air Unit 227 (not shown)

Identical to Unit 127 (Estimated instrument air requirement for the total complex is 750 lb-mol/hr; see Paragraph 3.5.7.)

4.5.8 Coal Handling Unit 228 (See SFD ZO-GEM-6909, Sub-Section 3.5.)

Gasifier well-sized coal production is identical to Base Case I. Because of the higher energy requirements in Base Case II, the quantity of boiler feed coal is increased to 406.9 Mlb/hr.

4.5.9 Ash Handling Unit 229 (See SFD ZO-GEM-6911, Sub-Section 3.5.)

Identical to Unit 129

4.5.10 BFW Preparation Unit 231

Deaerator (231/1)

Normal Capacity

	<u>Temp., °F</u>	<u>Flow, Mlb/hr</u>
MP condensate	479	346.6
LP condensate	297	1,628.5
Vacuum condensate	173*	2,221.1
Methanation Unit 212 process condensate	288	110.8
BFW make-up water	173*	2,099.4
Total	221	6,406.4

*After heating in 250-E-03.

Demineralizer (231/2)

Normal Capacity 2,090.4 Mlb/hr

4.5.11 Cooling Water Make-Up Preparation Unit 232

Normal Capacity

Mlb/hr

Treated fresh water	1,179.7
Treated water from Unit 235 and humidity condensed in Unit 221	1,961.9
Total	<u>3,141.6</u>

4.5.12 Cooling Water Tower Unit 233

Normal Capacity

GPM

Steam turbine condenser duty	174,300
Trim cooler duty	109,590
Total	<u>283,890</u>

4.5.13 Electric Power Generation Unit 234

Description

Generator 234-M-01 uses 48 psig, 378°F steam with
Condensing Turbine 234-T-01. Back Pressure Turbine
234-T-02 of Generator 234-M-02 is fed with two
superheated streams:

185 psig steam:	78.7 Mlb/hr
85 psig steam:	<u>777.1</u>
Total	<u>855.8</u>

Normal Capacity

MW

Steam Flow,
Mlb/hr

234-M-01	52.5	1,069.3
234-M-02	8.9	855.8

4.5.14 Waste Water Treatment Unit 235

Normal Capacity

Composition

Flow, Mlb/hr

Organic acids and phenols	8.9
Ammonia	1.3
Water containing some H ₂ S	2,093.4
Total	<u>2,103.6</u>

4.5.15 Flare and Blowdown Facilities 236 (not shown)

Normal Capacity

Blowdown water to holding pond 837.9 Mlb/hr

4.5.16 Storage 237 (not shown)

The following is a summary of the product, intermediate and chemical storage tank capacities which are not located within the process offsite units:

Product Storage (15 days)

Anhydrous ammonia	1,700 ton (refrigerated)
Sulfur	1,000 ton (covered)
Gasoline	224,000 Bbl (floating roof)
Diesel fuel	38,000 Bbl (cone roof)
Heavy fuel oil	11,000 Bbl (cone roof)
Propane LPG	18,500 Bbl (pressure)
Butane LPG	2,400 Bbl (pressure)
Alcohols	34,000 Bbl (floating roof)

Intermediate Storage

Hydrotreater Unit 251 feed (15 days)	22,000 Bbl (floating roof)
Hydrotreater Unit 253 feed (15 days)	188,000 Bbl (floating roof)
Reformer Unit 255 feed (5 days)	24,000 Bbl (floating roof)
Polymerization Unit 257 feed (2 days)	14,000 Bbl (pressure)

Chemical Storage

Diisopropyl ether	1,500 Bbl
Methanol	1,000 Bbl

4.5.17 Interconnecting Pipeway 238 (not shown)

Requirements are somewhat larger than in Base Case I because of the greater number of onsite units needed for F-T product upgrading.

4.5.18 Refrigeration Unit 241

Normal Capacity

<u>Temperature Level</u>	<u>Duty, MMBtu/hr</u>	
	<u>+ 32°F</u>	<u>- 45°F</u>
Purification Unit 205		
including ammonia storage	25.5	70.5
HC Recovery Unit 210	-	43.4
H ₂ Purification Unit 261	2.1	-
	<u>27.6</u>	<u>113.9</u>

4.5.19 Gasoline Blending Unit 270

Except for additional blending steams, identical to Unit 154.

4.5.20 F-T Catalyst Preparation Unit 271

Purpose of the Unit is to prepare from magnetite iron ore (Fe_3O_4) a promoted, reduced iron catalyst for the fluid-bed Fischer-Tropsch reactors.

Technology Used

The catalyst plant design is based on literature articles describing: (a) the manufacture of iron ammonia catalysts, (17), and (b) the Sasol catalyst plant (13). An overall general design has only been developed. Promoters used and dosages, concentrate specifications, specific operating conditions, etc., are confidential and, thus, unavailable. This unit has been added to the offsites because of the quantity of fused catalyst required and the problems of shipping the sensitive catalyst under a H₂ or inert gas atmosphere.

Process Description

Catalyst preparation consists of 3 steps: iron ore purification, catalyst fusion and size reduction, and catalyst reduction. Equipment contained in the iron ore purification area includes:

- Ore unloading and storage
- Rotary drier for moisture removal before grinding
- Rod mill for size reduction required by air table
- Air separation table for tailings removal
- Concentrate storage

Concentrate yield has been assumed to be 50 percent.

Equipment in the catalyst fusion and size reduction step includes:

- Electric resistance furnace for melting/intimate mixing of concentrate and promoters
- Gyratory crusher and ball mill for size reduction required by the fluid-bed reactor
- Air separator for elimination of fines and oversize particles
- Catalyst storage

Equipment required in catalyst reduction includes:

- Fluidized-bed reactor for contact of hydrogen and unreduced F-T catalyst
- Hydrogen heater
- Heat exchangers for condensing water product
- Hydrogen recycle compressor
- Recycle gas drier for moisture control
- Reduced catalyst storage under H_2 or inert gas atmosphere

Reduction takes place under elevated temperature and pressure.

Normal Capacity

Iron Ore	91 T/SD
Reduced Catalyst	30 T/SD

BASE CASE II
4.6 UTILITY REQUIREMENTS AND STEAM/BFW BALANCE

4.6.1 Utility Requirements

Utility requirements for the individual units can be found as follows:

<u>Utility</u>	<u>Source</u>
Fuel	Type and stream number: BFD ZO-GEM-6988 (Sub-Section 4.4) Quantity and composition: Material balance sheets (Appendix B)
Power	BFD ZO-GEM-6988
Cooling water	BFD ZO-GEM-6988
Boiler feed water	Steam/BFW Balance ZO-GEM-6986 (Sub-Section 4.6)
Steam	Steam/BFW Balance ZO-GEM-6986

In addition, offsite Sub-Section 4.5 contains utility information.

4.6.2 Steam/BFW Balance

The steam and boiler feed water flows are shown on Drawing ZO-GEM-6986. The steam/electric power balance is acceptable in accordance with the study bases. Only about 3 MW(e) is surplus in comparison with 268 MW of total electric and steam power required in the plant complex. In addition, all steam produced is utilized.

UNIT NO.	EQUIPMENT LIST	OPRS.	SPARE
231-C-01	STEAM TURBINE CONDENSATE VACUUM BLOWER	1	1
231-G-01	BFW MAKE UP PUMP	1	1
231-G-02	VACUUM CONDENSATE PUMP	1	1
231-G-03	LP BFW PUMP	2	1
231-G-04	200 PSIG BFW PUMP	1	1
231-G-05	HP BFW PUMP	1	1
231-G-06	HP BFW PUMP	2	2
234-M-01	MAIN ELECTRIC POWER GENERATOR	1	0
234-M-02	SECOND ELECTRIC POWER GENERATOR	1	0
201-T-01	LOCK GAS COMPRESSOR STEAM TURBINE	1	0
204-T-01	SHIFTED GAS BOOSTER COMPRESSOR STEAM TURBINE	1	0
205-T-01	RECTINOL VACUUM BLOWER HP STEAM TURBINE	4	0
205-T-02	RECTINOL VACUUM BLOWER LP STEAM TURBINE	4	0
205-T-03	RECTINOL ABSORBED HAIR FEED PUMP STEAM TURBINE	4	0
210-T-01	PUMP STEAM TURBINE	1	0
211-T-01	SEPTANE BASK OFFGAS COMPRESSOR STEAM TURBINE	1	0
211-T-02	PRESSURE SWING OFFGAS RECOMPRESSOR STEAM TURBINE	1	0
212-T-01	UNIT 211 HYDROGEN COMPRESSOR STEAM TURBINE	1	0
212-T-02	ETHANOLATION RECYCLE COMPRESSOR STEAM TURBINE	1	0
214-T-01	STEAM TURBINE	1	0
221-T-01	SWG COMPRESSOR STEAM TURBINE	1	0
221-T-02	OXYGEN COMPRESSOR STEAM TURBINE	1	0
221-T-03	AIR COMPRESSOR STEAM BASK	1	0
221-T-04	PRESSURE SWING RECYCLE COMPRESSOR STEAM TURBINE	1	0
221-T-05	BFW MAKE UP PUMP STEAM TURBINE	1	0
221-T-06	VACUUM CONDENSATE PUMP STEAM TURBINE	1	0
221-T-07	LP BFW PUMP STEAM TURBINE	1	0
221-T-08	200 PSIG BFW PUMP STEAM TURBINE	1	0
221-T-09	HP BFW PUMP STEAM TURBINE	2	0
221-T-10	HP BFW PUMP STEAM TURBINE	2	0
221-T-11	VACUUM CONDENSATION C.W. PUMP STEAM TURBINE	3	0
223-T-01	STEAM TURBINE	3	0
223-T-02	PROCESS C.W. PUMP STEAM TURBINE	1	0
224-T-01	MAIN ELECTRIC POWER GENERATOR	1	0
224-T-02	STEAM TURBINE	1	0
224-T-03	SECOND ELECTRIC POWER GENERATOR	1	0
224-T-04	STEAM TURBINE	1	0
224-T-05	HYDROLYSIS RECYCLE COMPRESSOR STEAM TURBINE	1	0
224-T-06	REFORMER RECYCLE COMPRESSOR STEAM TURBINE	1	0
225	BOILER	3	0
225	MAIN SUPERHEATER	3	0
224/2	200 PSIG STEAM SUPERHEATER	1	0
224/3	100 PSIG STEAM SUPERHEATER	1	0
224/4	DEMINERALIZER	3	0
224/5	DEMINERALIZER	1	0

Project	21.7.77	Rev	00-0865	LURGI
Client	Lurgi (Mitsubishi) GmbH			
Contract	Project (Main)			
Design	Steam/BFW Balance			
Drawn	MRO/ERDA STUDY			
Checked	EF-76-C-01-2-47			
Approved	BASE CASE II			

BASE CASE II

4.7 TRAIN PHILOSOPHY

The train philosophy for the various units was selected on the basis of providing a plant complex having an onstream factor of 92%. Consequently, parallel trains have been used for many units to give flexibility in both operation and maintenance. Also, because of the large size of the complex, some units must be divided into parallel trains. In the hydrocarbon processing area, however, intermediate storage is used to provide this flexibility.

In addition, all normally operating pumps, reciprocating compressors and conveyors are provided with an 100% spare.

Since the same train philosophy is used for Base Case II as is used in Base Case I, only the philosophy of the new units in Base Case II are discussed below.

Hydrocarbon Recovery Unit 210

Because of size limitation, two parallel trains are provided.

F-T Synthesis Unit 250

Using a conservative scale-up factor of 2 over current commercial fluid-bed reactors, four reactors are required. In addition, a spare reactor is provided for catalyst changing and maintenance. Size considerations also require multiple towers, compressors, and pumps; see equipment list on PFD ZO-GEM-6977 in Sub-Section 4.4.

F-T Product Upgrading Units 252 Through 262

These units are single trains as equipment sizes are small enough to be shop fabricated. The only exception is Catalyst Polymerization Unit 257 which is provided with two reactors for operating flexibility.

To achieve operating flexibility in the upgrading area, feedstock storage capacity has been provided for the following units:

Naphtha Hydrotreating Unit 251	-	15 days
F-T Product Hydrotreating Unit 253	-	15 days
Catalytic Reforming Unit 255	-	5 days
Catalytic Polymerization Unit 257	-	2 days

Offsite Units

The offsite units are designed for a minimum of two parallel trains. In cases where commercial-sized equipment is too large with two trains, the number of trains has been increased.

SECTION 5
SENSITIVITY CASES

The block flow diagram for each sensitivity case is placed at the end of its sub-section. The material balance of the new unit introduced in each sensitivity case is tabulated in Appendix C.

5.1. SENSITIVITY CASE I-A: METHANOL AND SNG COPRODUCTION

5.1.1 Material Balance

The overall material balance is shown below:

<u>Input</u>	<u>Mlb/hr</u>
Coal, as mined	2,277.8 ⁽¹⁾
Air	5,874.8
Water	3,094.2
	<u>11,246.8</u>
<u>Output</u>	<u>Mlb/hr</u>
Coal fines (excess)	132.2 ⁽²⁾
Ash	138.6
Products	865.5 ⁽³⁾
Byproducts	13.7
Blowdown water	707.8
Stack and evaporation losses	9,389.0
	<u>11,246.8</u>

(1) 27,334 T/SD

(2) 1,586 T/SD

(3) 47,880 FOE Bbl/SD @ 6.0 MM Btu/FOE Bbl

Overall plant consumptions per FOE barrel of product are:

Coal:	0.538 T
Water:	4.43 Bbl
Air:	1.47 T

Stack and evaporation losses amount to about 2.4 tons per FOE barrel of product.

5.1.2 Product Yields and Quality

Stream numbers are shown below for reference to
BFD ZO-GEM-7037.

(a) SNG (49)

<u>Quantity</u>	16,070 lb-mol/hr
	260.1 M lb/hr
	146.4 MMSCF/SD

Composition (only compounds greater than 0.1%
are listed.)

	<u>Percent</u>
Hydrogen	1.7
Methane	95.9
Carbon dioxide	0.5
Inerts (N ₂ + Ar)	1.9
	<u>100.0</u>

Other

Heat of combustion, Btu/SCF	
HHV	975
LHV	878
Carbon monoxide (0.1% max.)	0.01%
Water	0.01%
Sulfur	None

Compatibility Indexes (versus pure methane)

<u>Index</u>	<u>Calculated</u>	<u>Preferable</u>	<u>Objectionable</u>
Lifting, I _l	1.03	under 1.0	above 1.06
Flash-back, I _f	1.02	under 1.15	above 1.2
Yellow-tip, I _y	1.06	above 1.0	under 0.8

The interchangeability of the SNG product with pure methane is satisfactory and, essentially, unchanged from Base Case I.

(b) Methanol (37.1)

<u>Quantity</u>	589.9 M lb/hr 6,420 metric T/SD 50,840 Bbl/SD
-----------------	---

Impurities

	<u>PPM</u>
Light boiling compounds	1,200
Heavy boiling compounds	700
Water	1,500

Only a fuel grade methanol is prepared. For direct blending in gasoline, it would have to be dehydrated further to about 1,000 ppm of water.

(c) Naphtha (100)

<u>Quantity</u>	15.5 M lb/hr 1,315 Bbl/SD
-----------------	------------------------------

Quality

Gravity, °API	43.5
Octane, unleaded	
Research	93
Motor	84.5
Reid Vapor Pressure, lb	3.5

The hydrotreated naphtha is a good gasoline pool blending stock.

(d) Byproducts

<u>Sulfur</u> (29)	61 T/SD
--------------------	---------

<u>Anhydrous Ammonia</u> (17)	103 T/SD
-------------------------------	----------

<u>Excess Power</u>	- 6.55 MW(e) (required)
---------------------	-------------------------

(e) <u>Coal Fines</u> (2.1)	132.2 M lb/hr
-----------------------------	---------------

5.1.3 Thermal Efficiency

The overall plant thermal efficiency (HHV) is shown in Table 5.1.1. Deleting the Mobil methanol conversion technology increases the overall plant thermal efficiency from 62 to 66 percent. About half of this difference is caused by the inability to utilize the additional yield of low grade heat brought about by the methanol conversion to gasoline. (See Paragraph 5.1.6.)

5.1.4 Processing Description (BFD ZO-GEM-7037)

The sensitivity basis calls for the production of a fuel grade methanol. Consequently, Methanol Distillation Unit 115 is added for water removal only, with the trace non-methanol materials remaining. It is a two-stage tower system to minimize steam usage. Methanol Conversion Unit 150 and Units 152, 153 and 154 are deleted. The gasification and syngas purification areas are unchanged. Technology of the SNG train remains the same but the gas flow and composition differ slightly from those of Base Case I.

TABLE 5.1.1
SENSITIVITY CASE I-A - THERMAL EFFICIENCY

	<u>Quantity, Unit/SD</u>	<u>High Heating Value (HHV)</u>	<u>Total HHV, MM Btu/hr</u>	<u>Percent of Input</u>
Input				
Coal, DAF	18,286 T	12,720 Btu/lb	19,383	
<u>Fines (excess)</u>	<u>(1,061)</u>	" "	<u>(1,125)</u>	
Net Coal	17,225 T		18,258	
Power (required)	6.55 MW(e)	3,415 Btu/kwh*	<u>22</u>	
Total			18,280	
Output				
SNG	146.4 MMSCF	975 Btu/SCF	5,948	32.5
Methanol	7,080 T	9,724 Btu/lb	5,737	31.4
Naphtha	1,315 Bbl	5.2 MM Btu/Bbl	<u>285</u>	<u>1.5</u>
Sub-Total			11,970	65.4
Sulfur	61 T	3,780 Btu/lb	19	0.1
Ammonia	103 T	9,693 Btu/lb	<u>83</u>	<u>0.5</u>
Total			12,072	66.0*

*Direct thermal conversion used. If 10,000 Btu/kwh used, thermal efficiency becomes 65.8%.

5.1.5 Capacities of Process and Offsite Units

The following is a list of the process and offsite unit capacity ratios with Base Case I equaling 100 percent:

<u>Unit</u>	<u>Name</u>	<u>Ratio, %</u>
101 thru 109	Gasification and Syngas Purification	100
110	Methanol Synthesis	100
111	H ₂ Recovery	100
112	Methanation	100
113	CO ₂ Removal	98 (gas flow)
114	SNG Drying	98 (gas flow)
115	Methanol Distillation	New Unit
121	Oxygen Production	100
122/123	Boiler/Superheater	107 (heat fired)
124	Superheater	88 (heat fired)
125/126	Stackgas Precipitator & Clean-Up	94 (gas flow)
127	Instrument & Plant Air	100
128/129	Coal & Ash Handling	100
131/1 & 2	BFW Deaerator & Demineralizer	96
132	CW Make-Up Preparation	89
133	CW Towers	89
134	Electric Power Generation	76
135	Waste Water Treatment	82 (water flow)
136	Relief and Blow Down Facilities	88
137	Storage	(1)
138	Interconnecting Piping	(2)
141	Refrigeration	100
150	Methanol Conversion	0
151	Naphtha Hydrotreating	100
152	Fractionation	0
153	HF Alkylation	0
154	Gasoline Blending	0

- (1) Product storage deleted: Gasoline, Propane LPG and Mixed Butanes.

Intermediate storage deleted: Methanol

Product storage added:

Methanol (15 days)	840,000 Bbl (floating roof)
Hydrotreated Naphtha (15 days)	22,000 Bbl (floating roof)
Coal fines storage;	11,500 T (silo)

- (2) Interconnecting piping requirements are reduced due to the deletion of Units 150, 152, 153 and 154.

5.1.6 Steam Balance (as related to Base Case I)

The deletion of LP steam producer Unit 150 and the addition of LP steam consumer Unit 115 results in a decrease in the LP steam available for power generation. Consequently, insufficient power is generated, or purchase power is required, without a major modification to the steam balance. On the other hand, the deletion of Unit 150 (recycle compressor) reduces the HP steam, or boiler, requirement and results in a small increase of fine coal for sales.

In Base Case I, which has a surplus of low grade heat available for boiler feed water heating, all BFW heating is accomplished in the Methanol Conversion Unit 150. For this sensitivity case, the BFW heating duty of about 375 MM Btu/hr is spread among Units 103, 104, 110 and/or 112, all of which in Base Case I have low grade heat wasted in fan coolers. The utilization of the "waste" low grade heat has been a major problem in this study.

1) The steam/BFW circuit of the complex is shown simplified

GEM		Lurgi	
00-0865		Lurgi Mineralabsorbent GmbH	
BLOCK FLOW DIAGRAM		Z0-GEM-7037	
COAL TO METHANOL		MRO/ERDA STUDY	
CONVERSION		FF-76-C-01-227	
SENSITIVITY CASE 1-A			

5.2 SENSITIVITY CASE I-B:
GASOLINE ONLY PRODUCTION

5.2.1 Material Balance

The overall material balance is shown below:

<u>Input</u>	<u>M lb/hr</u>
Coal, as mined	2,286.6 ⁽¹⁾
Air	8,654.2
Water	3,419.3
	<u>14,360.1</u>
 <u>Output</u>	 <u>M lb/hr</u>
Ash	145.8
Products	448.5 ⁽²⁾
Byproducts	13.7
Blowdown water	971.2
Stack and evaporation losses	12,780.9
	<u>14,360.1</u>

(1) 27,439 T/SD

(2) 36,200 FOE Bbl/SD @ 6.0 MM Btu/FOE Bbl

Overall plant consumptions per FOE barrel of product are:

Coal:	0.758 T
Water:	6.48 Bbl
Air:	2.87 T

Stack and evaporation losses amount to about 4.2 tons per FOE barrel of product.

Although methane reforming increases the C₃ plus hydrocarbon yield from 25,805 to 43,895 Bbl/SD, or 70 percent, the total hydrocarbon yield shrinks from 45,550 to 36,200 FOE Bbl/SD, or 20 percent. This decrease results in the larger overall plant consumptions shown above as compared to those for Base Case I.

5.2.2 Product Yields and Quality

Stream numbers are shown below for reference to
BFD ZO-GEM-7036.

(a) Gasoline (50)

<u>Quantity</u>	37,430 Bbl/SD 397.2 M lb/hr
-----------------	--------------------------------

Blending and Properties

The 10 RVP gasoline consisting of the same components as in Base Case I, essentially, has the same properties shown in Table 3.2.1 and, therefore, is of excellent quality.

(b) Mixed Butanes (103)

<u>Quantity</u>	3,800 Bbl/SD 31.6 M lb/hr
-----------------	------------------------------

Properties

Estimated properties are the same as for the mixed butanes obtained in Base Case I. (See Table 3.2.3.)

(c) Propane LPG (102)

<u>Quantity</u>	2,665 Bbl/SD 19.7 M lb/hr
-----------------	------------------------------

Properties

Estimated properties are the same as for the propane LPG obtained in Base Case I. (See Table 3.2.3.)

(d) Byproducts

<u>Sulfur</u> (29)	61 T/SD
<u>Ammonia</u> (17)	103 T/SD
<u>Excess Power</u>	1.38 MW(e)

5.2.3 Thermal Efficiency

The overall plant thermal efficiency (HHV) is shown in Table 5.2.1. The addition of methane reforming for the elimination of SNG production reduces the overall plant thermal efficiency by about 15 percent. Only about 61% of the HHV energy in the SNG product of Base Case I is realized in additional C₃ plus liquid hydrocarbon products. In addition, the fine coal usage is slightly higher than in Base Case I.

5.2.4 Processing Description (BFD ZO-GEM-7036)

To maximize the production of gasoline - deletion of the SNG coproduct - Autothermal Reforming Unit 116 is added to convert the purge (38) and expansion (35) gases from Methanol Synthesis Unit 110 and the light gases (57) from Fractionation Unit 152 into an additional synthesis gas feed (42) for Unit 110. The SNG train units are deleted.

In the autothermal reforming process, direct partial combustion of the feed gases with oxygen produces the required heat of reaction. This technique, as opposed to tubes-in-firebox heating in the conventional steam reforming process, yields the maximum possible syngas.

Because of the nitrogen inerts in the oxygen feed and from the coal gasification itself, not all of the methanol unit purge gas can be recycled to Autoreforming Unit 116. A slipstream (39) of about 9% of the methanol unit purge gas (38) is required to maintain the inerts at a reasonable level in Methanol Synthesis Unit 110. This slipstream is used to fire Boiler 122. A partial slipstream reduction, thereby increasing somewhat the gasoline yield, could be achieved by the use of a higher purity oxygen plant. Because of the nitrogen yield in the gasification step - not all the coal nitrogen is converted into ammonia -, however, the slipstream cannot become zero.

TABLE 5.2.1
SENSITIVITY CASE I-B - THERMAL EFFICIENCY

	<u>Quantity, Unit/SD</u>	<u>High Heating Value (HHV)</u>	<u>Total HHV, MMBtu/hr</u>	<u>Percent of Input</u>
Input				
Coal, DAF	18,357 T	12,720 Btu/lb	19,458	
Output				
C ₃ LPG	2,665 Bbl	3.816 MMBtu/Bbl	424	2.2
C ₄ LPG	3,800 Bbl	4.191 MMBtu/Bbl	664	3.4
10 RVP Gasoline	37,430 Bbl	5.105 MMBtu/Bbl	<u>7,962</u>	<u>40.9</u>
Sub-Total			9,050	46.5
Sulfur	61 T	3,780 Btu/lb	19	0.1
Ammonia	103 T	9,693 Btu/lb	83	0.5
Power (excess)	1.38 MW(e)	3,415 Btu/kwh*	<u>5</u>	<u>-</u>
Total			9,157	47.1

*Direct thermal conversion used.

5.2.5 Capacities of Process and Offsite Units

The following is a list of the process and offsite unit capacity ratios with Base Case I equaling 100%.

<u>Unit</u>	<u>Name</u>	<u>Ratio, %</u>
101	Gasification	100
	Lock Gas Compressors	107(1)
102	Raw Gas Shift	100
103	Raw Gas Cooling	100
104	Shifted Gas Cooling	100
105	Gas Purification	102(2)
106	Sulfur Recovery	100(3)
107	Gas Liquor Separation	100
108	Phenol Recovery	100
109	Ammonia Recovery	100
110	Methanol Synthesis	173 (methanol(4) yield)
111	H ₂ Recovery	100
112	Methanation	0
113	CO ₂ Removal	0
114	SNG Drying	0
116	Autothermal Reforming	New Unit
121	Oxygen Production	153(5)
122	Boiler	124 (steam flow)
123	Main Superheater	126 (steam flow)
124	Superheater	115 (heat fired)
125/126	Stackgas Precipitation & Clean-Up	120 (gas flow)
127	Instrument & Plant Air	100
128/129	Coal & Ash Handling	100
131/1 & 2	BFW Deaerator & Demineralizer	96
132	CW Make-Up Preparation	124
133	CW Towers	124
134	Electric Power Generation	112
135	Waste Water Treatment	114 (water flow)
136	Relief and Blow Down Facilities	100
137	Storage	(6)
138	Interconnecting Piping	100 (differences negligible)
141	Refrigeration	100
150	Methanol Conversion	173 (methanol(7) feed)
151	Naphtha Hydrotreating	100
152	Fractionation	173
153	HF Alkylation	171
154	Gasoline Blending	170

- (1) Increased syngas compressor (110-C-01) losses from larger syngas feed to Methanol Synthesis Unit 110.
- (2) A lower CO₂ content, clean synthesis gas is required from Unit 105 to maintain the CO₂ specification for Methanol Synthesis Unit 110 with the addition of synthesis gas from Autothermal Reforming Unit 116. Investment is estimated to be 2% higher.
- (3) Feed to Unit 106 contains slightly more CO₂ than in Base Case I. Effect on investment is neglected.
- (4) The H₂ and CO conversions are higher than those for Base Case I because of the lower methane content in the syngas.
- (5) The oxygen feed pressure required by Autothermal Reforming Unit 116 is the same as required by Gasification Unit 101.
- (6) Storage capacity for crude methanol, C₃ LPG, C₄ LPG and 10 RVP gasoline is increased by 70%.
- (7) Two parallel reactor trains are required.

5.2.6 Steam Balance (as related to Base Case I)

The HP steam (Boiler/Superheater Unit 122/123) requirement is increased to meet the needs of the larger oxygen plant, methanol synthesis unit and methanol conversion unit compressors. Additional boiler firing is with the methanol/autothermal reforming slipstream gas for inerts control and coal fines. The increased steam production in Methanol Synthesis Unit 110 reduces the amount of 550 psig steam generation from the HP steam system for feed to Gasification Unit 101. Electricity production is slightly higher due to the increase yield of LP steam from Methanol Conversion Unit 150.

1) The steam/BFW circuit of the complex is shown simplified

GEM		Lurgi	
00-0865		Lurgi Process Development GmbH	
BLOCK FLOW DIAGRAM		Z0-GEM-7036	
COAL TO GASOLINE VIA		MPC/EPDA STUDY	
METH'L WITHOUT SNG PROD		EF-75-C-01-104	
SENSITIVITY CASE 1-B			

5.3 SENSITIVITY CASE I-C:
MOBIL FLUID-BED TECHNOLOGY

5.3.1 Material Balance

The overall material balance is shown below:

<u>Input</u>	<u>M lb/hr</u>
Coal, as mined	2,277.8 ⁽¹⁾
Air	5,862.6
Water	2,981.5
	<u>11,121.9</u>
<u>Output</u>	<u>M lb/hr</u>
Coal fines (excess)	137.8 ⁽²⁾
Ash	138.3
Products	529.9 ⁽³⁾
Byproducts	13.7
Blowdown water	781.7
Stack and evaporation losses	9,520.5
	<u>11,121.9</u>

(1) 27,334 T/SD

(2) 1,654 T/SD

(3) 45,560 FOE Bbl/SD @ 6.0 MM Btu/FOE Bbl

Overall plant consumptions per FOE barrel of product are:

Coal:	0.564 T
Water:	4.49 Bbl
Air:	1.54 T

Stack and evaporation losses amount to about 2.5 tons per FOE barrel of product. The barrels of methanol feed required to yield a barrel of 10 RVP gasoline (without the Lurgi gasifier naphtha) is 2.33.

The overall plant performance is slightly superior when shifting to fluid-bed operation.

5.3.2 Product Yields and Quality

Stream numbers are shown below for reference to
BFD ZO-GEM-7038.

(a) SNG (49)

<u>Quantity</u>	16,756 lb-mol/hr
	270.7 M lb/hr
	152.6 MMSCF/SD

Composition (Only compounds greater than 0.1%
are listed.)

	<u>Percent</u>
Hydrogen	1.8
Methane	95.9
Carbon dioxide	0.5
Inerts (N ₂ + Ar)	1.8
	<u>100.0</u>

Other

Heat of combustion	
HHV	976 Btu/SCF
LHV	878 Btu/SCF
Carbon monoxide (0.1% max)	0.02%
Water	0.01%
Sulfur	None

Compatibility Indexes (versus pure methane)

<u>Index</u>	<u>Calculated</u>	<u>Preferable</u>	<u>Objectionable</u>
Lifting, I _l	1.03	under 1.0	above 1.06
Flash-back, I _f	1.02	under 1.15	above 1.2
Yellow-tip, I _y	1.06	above 1.0	under 0.8

The SNG product is of satisfactory quality and is
interchangeable with pure methane.

Its composition and properties differ only slightly from
those for the SNG produced in Base Case I.

(b) Gasoline (50)

Quantity

23,065 Bbl/SD
246.1 Mlb/hr

Blending

<u>Component</u>	<u>Wt. %</u>
Mixed Butanes	2.3
Alkylate	24.6
Stabilized gasoline	66.8
Hydrotreated gasifier naphtha	6.3
	100.0

Properties

Estimated properties of the unleaded gasoline are presented in Table 5.3.1. (Gasoline specifications are listed in Table 3.2.2.)

Alkylation of the C₃ and C₄ olefins reduces the iso-butane to such a low net yield that the Reid Vapor Pressure of the blended gasoline is only 9.4 psig, falling slightly short of the 10 RVP target. To make 10 RVP, an additional 285 Bbl/SD of n-butane is estimated to be required.

The fluid-bed operation yields less stabilized gasoline and more alkylation feed stock than the fixed-bed operation, resulting in a higher final gasoline yield, 23,065 vs. 22,045 B/SD, and a more paraffinic final gasoline, 63 vs. 51%. The motor octane of the fluid-bed gasoline is two numbers higher. In addition, its durene content is much lower at 2.3 wt. %. Thus, the fluid-bed gasoline is of excellent quality and slightly superior to the Base Case I, or fixed-bed, gasoline.

(c) Mixed Butanes (103)

Quantity

0 Mlb/hr

All the butanes are used in alkylation and pressuring the gasoline.

TABLE 5.3.1
ESTIMATED GASOLINE PROPERTIES FOR
SENSITIVITY CASE I-C: MOBIL FLUID-BED TECHNOLOGY

	<u>Estimated Unloaded Properties</u>
Gravity, °API	61.8
Octane Numbers	
Research	93
Motor	85
(Research + Motor)/2	89
Volatility	
Reid Vapor Pressure, lb	9.4*
Distillation, °F	
IBP	87
10%	111
30%	152
50%	216
70%	259
90%	341
EP	404
V/L Ratio (=20), °F	@ 132
Sulfur, wt. %	nil
Composition, vol. %	
Paraffins	63
Olefins	6
Naphthenes	5
Aromatics	26
Durene Content, wt. %	2.3
Molecular Weight	96

* Target = 10.0 RVP

(d) Propane LPG (102)

Quantity

1,790 B/SD
13.1 Mlb/hr

Properties

Estimated properties are as follows:

Vapor pressure at 100°F, psig	180*
Butane and heavier, Vol. %	1.2

*Commercial specifications are given in Table 3.2.3.

This product is a satisfactory commercial propane fuel.

(e) Byproducts

Sulfur (29)

61 T/SD

Ammonia (17)

103 T/SD

Excess Power

-0.08 MW(e) (required)

(f) Coal_Fines (2.1)

137.8 Mlb/hr

(g) Comparison with Base Case I

Methanol conversion in a fluid-bed yields a higher percentage of light hydrocarbons. Consequently, more product ends up in SNG, thereby increasing the SNG yield from 53.3 to 54.5 percent of the hydrocarbon thermal yield. On the other hand, the C₃ and C₄ mix of iso-butane and olefins enables greater use of alkylation to increase the percentage of gasoline from 41.2 to 43.0 percent. The LPG yield is reduced from 5.5 to 2.5 percent.

The total weight of hydrocarbon yield is slightly lower by 1.0 Mlb/hr, principally, due to the disposition of the offgases from methanol conversion. In fixed-bed operation, they are blended into the SNG train after methanation. Because of the increased H₂ and CO content in fluid-bed operation, they are added before methanation. Consequently, water is formed and a SNG weight loss occurs.

TABLE 5.3.2
SENSITIVITY CASE I-C - THERMAL EFFICIENCY

	<u>Quantity, Unit/SD</u>	<u>High Heating Value (HHV)</u>	<u>Total HHV, MMBtu/hr</u>	<u>Percent of Input</u>
Input				
Coal, DAF	18,286 T	12,720 Btu/lb	19,383	
<u>Fines (excess)</u>	<u>(1,106 T)</u>	" "	<u>(1,172)</u>	
Net Coal	17,180	" "	18,211	
Power (required)	0.08 MW(e)	3,415 Btu/kwh*	<u>nil</u>	
Total			18,211	
Output				
SNG	152.6 MMSCF	976 Btu/SCF	6,206	34.1
C ₃ LPG	1,790 Bbl	3.793 MMBtu/Bbl	283	1.5
C ₄ LPG	0	-	-	
10 RVP Gasoline	23,065 Bbl	5.100 MMBtu/Bbl	<u>4,901</u>	<u>26.9</u>
Sub-Total			11,390	62.5
Sulfur	61 T	3,780 Btu/lb	19	0.1
Ammonia	103 T	9,693 Btu/lb	<u>83</u>	<u>0.5</u>
Total			11,492	63.1

*Direct thermal conversion used.

5.3.3 Thermal Efficiency

The overall plant thermal efficiency (HHV) is shown in Table 5.3.2. The fluid-bed methanol conversion technology is slightly more efficient, about 1%, than the fixed-bed technology. It arises not only because of marginally improved yields, but because of less HP steam required - large fixed-bed recycle compressor eliminated - and 550 psig steam generation in the fluid-bed unit.

5.3.4 Processing Description (BFD ZO-GEM-7038)

The introduction of a fluid-bed reactor results in only very minor changes in the block flow diagram, i.e., (1) elimination of fuel gas stream 101 to Unit 150, (2) shifting of Unit 150 steam yield stream to the 571 psig steam header, (3) slightly different handling of the light gas yield in and out of Units 150 and 152, and (4) elimination of stack gas stream 108.

One large fluid-bed reactor, instead of six fixed-bed reactors, is the nucleus of Methanol Conversion Unit 150. The projected operating temperature and pressure are 750°F and 40 psig, respectively. The conceptual reactor design used in this study includes internal baffles and an outside catalyst flow leg to insure intimate catalyst/methanol mixing. To absorb the heat of reaction, the feed is 100 percent liquid. Current on-going pilot plant studies are indicating that for better operation, vapor feed should be used with temperature control via internal steam generation coils. A small recycle stream may also be required. A slipstream of catalyst is continuously removed and regenerated in a separate regeneration vessel. For catalyst activity control, the coke level of the regenerated catalyst is maintained at about 12 wt. %. The reactor effluent cooling system is somewhat more complex for catalyst fines recovery and return to the reactor. The yields, summarized in Appendix C, are based on work completed under DOE Contract No. E(49-18)-1773(2). The development of the fluid-bed technology continues under pilot plant DOE Contract No. EX-76-C-01-2490.

In Fractionation Unit 152, a deethanizer absorber is employed in place of the Base Case I high-pressure deethanizer tower. The light gas overhead is split into a recycle stream to Unit 150 for increasing the C₃ plus yield via additional ethylene reaction, a fuel gas stream and an SNG stream. In addition, a rich oil tower is required to produce a lean stream for recycle. Other than size shifts, Alkylation Unit 153 and Gasoline Blending Unit 154 are identical to those used in Base Case I.

5.3.5 Capacities of Process and Offsite Units

The following is a list of the process and offsite unit capacity ratios with Base Case I equaling 100 percent.

<u>Unit</u>	<u>Name</u>	<u>Ratio, %</u>
101 thru 111	Gasification et al	100
112	Methanation	100
113	CO ₂ Removal	103 (gas flow)
114	SNG Drying	103 (gas flow)
121	Oxygen Production	100
122/123	Boiler/Superheater	92
124	Superheater	104 (heat fired)
125/126	Stackgas Precipitator & Clean-Up	95 (gas flow)
127	Instrument & Plant Air	100
128/129	Coal & Ash Handling	100
131/1	BFW Deaerator	97
131/2	BFW Demineralizer	100
132	CW Make-Up Preparation	96
133	CW Towers	96
134	Electric Power Generation	94
135	Waste Water	100
136	Relief and Blow Down Facilities	97
137	Storage	*
138	Interconnecting Piping	100
141	Refrigeration	100
150	Methanol Conversion (Fluid-Bed)	100 (methanol feed)
151	Naphtha Hydrotreating	100
152	Fractionation	100 (3 tower system)
153	Alkylation	800 (alkylate yield)
154	Gasoline Blending	105

*Storage adjustments:

C ₃ LPG (15 days)	29,000 Bbl (pressure)
C ₄ LPG	Delete
Gasoline (15 days)	378,000 Bbl (floating roof)

All other storage remains the same as for Base Case I.

5.3.6 Steam Balance (as related to Base Case I)

The elimination of the large recycle compressor for the fixed-bed Unit 150 and the generation of 550 psig steam in the fluid-bed Unit 150 reduces slightly Boiler/Superheater Units 122/123 heat duties, thereby releasing more coal fines for sales. The elimination of LP steam generation in Unit 150, however, reduces the amount of electric power generated.

1) The steam/BFW circuit of the complex is shown simplified

GEM		LURGI	
00-0855		Large Process/Production Plant	
BLOCK FLOW DIAGRAM		20-GEM-7038	
COAL TO GASOLINE CONV.		PROCESSED STUDY	
(FLUID BED) VIA METHANOL		EF-76-C-01-2447	
SENSITIVITY CASE I-C		Page 1 of 1	

5.4 SENSITIVITY CASE I-D: SECOND GENERATION GASIFIER

As stipulated in the contract, the data for the second generation gasifier sensitivity case were to have been provided by DOE. Subsequently, DOE advised that they were having difficulty in obtaining the necessary proprietary gasifier data. As a consequence, MRDC contacted Davy Powergas, Inc. of Lakeland, Florida, to determine if they would provide data on the Winkler pressure fluid-bed gasifier, a second generation gasifier currently under development in Germany, for use in a sensitivity case. Davy Powergas offered to provide Winkler data without any fee. However, Davy requested to review the portions of the report dealing with the Winkler process. Davy Powergas subsequently provided an equipment list, process flow diagrams, gasifier feed data, syn gas yield and composition, utility requirements, and an investment estimate.

A sensitivity case has been developed around these data. However, in developing the case, it became evident that it would be unrealistic to integrate the Winkler gasifier into a plant complex based on the yields and process parameters of the Lurgi gasifier. Briefly, problem areas include syngas of different composition, smaller purge gas streams, large production of char, and the different gasifier temperature and pressure levels. Integrating these process variations directly into the Lurgi process scheme did not allow proper optimization of equipment, purge gas utilization, and waste heat recovery. A sensitivity case using the advanced gasifier requires developing the plant complex in the same detail as for the Lurgi base case. Such a case is entirely beyond the current scope of work. A Davy Powergas review of the work confirms the problem areas discussed above, and makes concrete suggestions to improve the process. Consequently, the second generation sensitivity will not be reported here.

The sensitivity for using second generation gasifiers should be done in a separate study and in the same depth that this study was performed. The study could cover the high pressure Winkler, the Texaco Partial Oxidation gasifier, the slagging Lurgi, and other promising gasifiers and would provide guidance toward the development of the entire coal-to-gasoline process.

5.5 SENSITIVITY CASE II-A:
MOBIL DIRECT SYNGAS CONVERSION

5.5.1 Material Balance

The overall material balance is shown below:

<u>Input</u>	<u>Mlb/hr</u>
Coal, as mined	2,277.8 ⁽¹⁾
Air	6,917.5
Water	3,306.2
Other	3.2
	<u>12,504.7</u>

<u>Output</u>	<u>Mlb/hr</u>
Coal Fines (excess)	28.0 ⁽²⁾
Ash	144.4
Products	522.0 ⁽³⁾
Byproducts	13.7
Blowdown water	840.3
Stack and evaporation losses	10,953.8
Other	2.5
	<u>12,504.7</u>

(1) 27,334 T/SD

(2) 336 T/SD

(3) 45,650 FOE Bbl/SD @ 6.0 MM Btu/FOE Bbl

Overall plant consumptions per FOE barrel of product are:

Coal:	0.591 T
Water:	4.96 Bbl
Air:	1.82 T

Stack and evaporation losses amount to about 2.9 tons per FOE barrel of product.

Product yield, overall consumptions and stack emissions show an improvement over those obtained for the Fischer-Tropsch technology in Base Case II.

5.5.2 Product Yields and Quality

Stream numbers are given below for reference to
BFD ZO-GEM-7040.

(a) SNG (49)

<u>Quantity</u>	19,530 lb-mol/hr
	323.0 Mlb/hr
	177.9 MMSCF/SD

Composition (Only compounds greater than 0.1% are listed.)

	<u>Percent</u>
Hydrogen	1.5
Methane	93.9
Ethene	0.2
Ethane	2.1
Propene	0.1
Propane	0.2
Carbon dioxide	0.4
Inerts (N ₂ + Ar)	1.6
	<u>100.0</u>

Other

Heat of combustion	
HHV	1,002 Btu/SCF
LHV	902 Btu/SCF
Carbon monoxide (0.1% max.)	0.1%
Water	0.01%
Sulfur	None

Compatibility Indexes (versus pure methane)

<u>Index</u>	<u>Calculated</u>	<u>Preferable</u>	<u>Objectionable</u>
Lifting, I _l	1.02	Under 1.0	Above 1.06
Flash-back, I _f	1.03	Under 1.15	Above 1.2
Yellow-tip, I _y	1.02	Above 1.0	Under 0.8

The SNG product is of satisfactory quality and is compatible with pure methane. Its composition approaches the composition of the Base Case I SNG in that the methane content is over 90% and the olefin content is a low 0.3%.

(b) Gasoline (50)

Quantity

17,485 B/SD
186.6 Mlb/hr

Blending

<u>Component</u>	<u>Wt. %</u>
Mixed butanes	2.5
Alkylate	5.1
Hydrotreated gasifier naphtha	8.2
Stabilized gasoline	84.2
	<u>100.0</u>

Properties

Estimated properties of the unleaded gasoline are presented in Table 5.5.1 (Gasoline specifications are listed in Table 3.2.2).

The gasoline formulation is quite similar to that in Base Case I. Its composition and properties are approximately in between those of the base case gasolines and make a satisfactory commercial motor fuel.

(c) Mixed Butanes (139)

Quantity

0 Mlb/hr

All the butanes are used in alkylation and pressuring the gasoline.

(d) Propane LPG (138)

Quantity

1,675 Bbl/SD
12.4 Mlb/hr

Properties

Estimated properties are as follows:

Vapor pressure at 100°F, psia	178*
Butane and heavier, vol. %	0.7

*Commercial specifications are given in Table 3.2.3.

This product is a satisfactory commercial propane fuel.

TABLE 5.5.1
ESTIMATED GASOLINE PROPERTIES FOR
SENSITIVITY CASE II-A: MOBIL DIRECT ROUTE

	<u>Estimated Unleaded Properties</u>
Gravity, °API	61.8
Octane Numbers	
Research	92.5
Motor	82.5
(Research + Motor)/2	87.5
Volatility	
Reid Vapor Pressure, lb	10.2
V/L Ratio (=20), °F	estimated satisfactory
Sulfur, wt. %	nil
Composition, vol. %	
Paraffins	54
Olefins	18
Naphthanes	7
Aromatics	21
Molecular Weight	93

(e) Byproducts

<u>Sulfur</u> (29)	61 T/SD
<u>Ammonia</u> (17)	103 T/SD
<u>Excess Power</u>	-1.01 MW(e) (required)

(f) Coal_Fines (2.1) 28.0 Mlb/hr

(g) Comparison_with_Base_Cases

Converting to a FOE barrel, the direct route has a liquid product yield slightly greater than that for the Fischer-Tropsch technology, 45,650 vs. 44,950 Bbl/SD. Product selectivity, however, is improved as shown below by a comparison of the hydrocarbon product distributions on a thermal, or Btu, basis.

Hydrocarbon Product Thermal Yields

<u>Case</u>	<u>I</u>	<u>II</u>	<u>II-A</u>
SNG, %	53	65	65
Gasoline, %	41	25	33
Gasoline, Bbl/SD	22,045	13,580	17,485
Other, %	6	10	2
Total Product, % of input	61.5	57.0	59.6

The direct conversion technology results in a 3 percent improvement in the hydrocarbon yield over Base Case II, but it fails to achieve the Base Case I yield by 2 percent. Although there is a significant increase in the gasoline yield, about 28%, for the sensitivity case over its base case, it falls short by about 28 percent (4,560 Bbl/SD) of equaling the yield for Base Case I. Although more selective in yielding gasoline, the direct conversion technology, however, does not show a reduction in the SNG yield.

5.5.3 Thermal Efficiency

The overall plant thermal efficiency (HHV) is shown in Table 5.5.2. The direct syngas conversion technology yields a 2 to 3 percent efficiency improvement over the F-T technology. It falls about half-way between the two base cases. This improvement is realized, in 70/30 ratio, by a lower coal input and a higher product yield.

TABLE 5.5.2
SENSITIVITY CASE II-A - THERMAL EFFICIENCY

	<u>Quantity, Unit/SD</u>	<u>High Heating Value (HHV)</u>	<u>Total HHV, MMBtu/hr</u>	<u>Percent of Input</u>
Input				
Coal, DAF	18,286 T	12,720 Btu/lb	19,383	
<u>Fines (excess)</u>	<u>(225 T)</u>	<u>" "</u>	<u>(238)</u>	
Net Coal	18,061 T		19,145	
Methanol	8 T	9,724 Btu/lb	7	
Power (required)	1.01 MW(e)	3,415 Btu/kwh	3	
Total			<u>19,155</u>	
Output				
SNG	177.9 MMSCF	1,002 Btu/SCF	7,427	38.8
C ₃ LPG	1,675 Bbl	3,822 MMBtu/Bbl	267	1.4
C ₄ LPG	0	--	--	--
10 RVP Gasoline	17,485 Bbl	5,105 MMBtu/Bbl	3,719	19.4
Sub-Total			<u>11,413</u>	<u>59.6</u>
Sulfur	61 T	3,780 Btu/lb	19	0.1
Ammonia	103 T	9,693 Btu/lb	83	0.4
Total			<u>11,515</u>	<u>60.1</u>

5.5.4 Processing Description (BFD ZO-GEM-7040)

The development of the direct syngas conversion technology is currently in progress under the DOE Contract EX-76-C-01-2276. The product selectivity and properties used are extrapolations, or targets, of recent (July, 1977) laboratory data. The resultant yields are given in Appendix C. Also, there is insufficient data upon which to make a scoping reactor design. Consequently, the Fischer-Tropsch reactor, Unit 250, of Base Case II is used directly. The product recovery and upgrading units are modified to fit the target yields and properties.

As with the F-T technology, a large gas stream containing unreacted H₂ and CO, inerts, methane and conversion products is sent from Conversion Unit 250 to HC Recovery Unit 210. In this unit, SNG precursors and liquid hydrocarbons are separated, with the latter going to Fractionation Unit 252.

Since the reaction product has been estimated to be very similar to the one obtained in the methanol conversion technology, product recovery and upgrading is much simplified over Base Case II. Thus, only Fractionation Unit 252 and HF Alkylation Unit 258 have been retained and the following units have been deleted:

- F-T Product Hydrotreating Unit 253
- Hydrotreated Product Fractionation Unit 254
- Catalytic Reforming Unit 255
- C₅/C₆ Isomerization Unit 256
- Catalytic Polymerization Unit 257
- Poly Gasoline Hydrogeneration Unit 259
- Light Ends Recovery Unit 260
- H₂ Purification Unit 261
- Alcohol Recovery Unit 262

Fractionation Unit 252 is a 3-tower system: deethanizer absorber, debutanizer and lean oil regenerator. The light gases are used as process heater fuel with the excess being added to the SNG train. The debutanizer overhead is the feed to Alkylation Unit 258. Other than size shifts, Unit 258 and Gasoline Blending Unit 270 are identical to those used in Base Case I. Catalyst Preparation Unit 271 has been eliminated.

5.5.5 Capacities of Process and Offsite Units

The following is a list of the process and offsite capacity ratios with Base Case II equaling 100 percent.

Unit	Name	Ratio, %
201 thru 209	Gasification et al	100
210	Hydrocarbon Recovery	120
211	H ₂ Recovery	100
212	Methanation	103
213	CO ₂ Removal	100
214	SNG Drying & Compression	103 (gas flow)
221	Oxygen Production	100
222/223	Boiler/Superheater	88 (steam flow)
224/1 & 2	Superheater	517
225/226	Stackgas Precipitator & Clean-Up	91 (gas flow)
227	Instrument & Plant Air	100
228/229	Coal & Ash Handling	100(2)
231/1 & 2	BFW Deaerator & Demineralizer	100
232	CW Make-Up Preparation	100
233	CW Towers	100
234	Electric Power Generation	99
235	Waste Water Treatment	97
236	Relief and Blow Down	100
237	Facilities	(1)
238	Interconnecting Piping	100
241	Refriegeration Unit	107
250	Conversion Reactor	100 (gas flow)
251	Naphtha Hydrotreating	100
252	Product Fractionation	New Design
253 thru 257	Product Upgrading	0
258	HF Alkylation	126 (alkylate)(2)
259 thru 262	Product Upgrading	0
270	Gasoline Blending	79 (2)
271	Catalyst Preparation	0
(1)	Complete storage is as follows:	

Product (15 days)	
Ammonia	1,700 T (refrigerated)
Sulfur	1,000 T (covered)
Gasoline	288,000 Bbl (floating roof)
C ₃ LPG	20,000 Bbl (pressure)
Intermediate (15 days)	
Unit 251 feed	22,000 Bbl (floating roof)

Chemical	
Diisopropyl ether	1,500 Bbl
Methanol	2,300 Bbl

(2) Basis is 100% Base Case I.

5.5.6 Steam Balance (as related to Base Case II)

The deletion of the MP steam users (F-T upgrading units) reduces the need for MP steam generation from HP steam. Thus, the boiler size is significantly reduced releasing fines for sales.

Project	Case	Name	GEM	LURGI
00-0865	00-0865	00-0865	00-0865	00-0865
BLOCK FLOW DIAGRAM				
COAL TO GASOLINE CONV.				
VIA DIRECT ROUTE				
SENSITIVITY CASE II A				
Z0-GEM-7040				
Drawing No.				
Scale				
Date				
By				
Checked by				
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SECTION 6
COST ESTIMATION

6.1 INVESTMENT

6.1.1 Bases

The bases for the investment estimates are outlined below:

- (a) Pricing for equipment and material reflect October, 1977 costs. No escalation beyond October has been included.
- (b) Direct labor unit rates used are based on Gulf Coast 1970, adjusted to represent the Gillette, Wyoming area for October, 1977. An overall weighted average wage rate of \$11.38 per manhour, excluding fringes and burden, for a 40 hour week is used. The Wyoming labor productivity factor for October, 1977 has been estimated to be 1.63 times Gulf Coast 1970. No escalation beyond October has been included.
- (c) Field indirect costs have been based on 110 percent of direct field labor costs.
- (d) Freight on equipment and materials has been estimated at 6 percent.
- (e) Home office (engineering + design) costs have been set at 11.5 percent of the field construction costs.
- (f) Camp costs for construction labor equal 6 percent of the contractor cost. From experience, monies are required to establish suitable living conditions for the construction workers in remote, non-industrial areas, such as the Wyoming coal fields.
- (g) Overtime premium of double-time pay for 14 hours per week is included. In order to attract the necessary large numbers of construction workers to this isolated site, a 54 hour work week has been projected. (Depending upon the construction environment assumed, the base work week could range from 45 to 60 hours.)
- (h) Estimating allowance at 15 percent is used. This allowance is required to obtain an 80 percent probability for the cost estimate to be within + 20 percent of the actual cost of the project.
- (i) Sales tax of 3 percent of the material cost is included.

- (j) Other items included are: contractor fee, capitalized spare parts, process unit data logger, project management cost, builder's all-risk insurance premium, catalyst and chemical cost, and paid-up royalties.

An allowance of two million dollars for environmental studies and impact reports has been added to the first two years of the project.

Catalysts estimated to have lives greater than 2 years are capitalized.

The following items have been excluded:

- a. Land or land rights
- b. Escalation
- c. Special foundation conditions
- d. Tie-ins outside of plant area required for road, railroad, water supply, power and pipelines
- e. Catalyst and royalties for Mobil methanol conversion and Mobil direct route technologies
- f. Royalty for Fischer-Tropsch technology

In our opinion, the bases adopted above will yield more complete and realistic plant investment estimates than those delineated in the Gas Cost Guidelines.

6.1.2 Base Cases

The total plant investments for the base cases are estimated to be:

	<u>MM \$</u>
Base Case I -	1,732
Base Case II-	1,887

The difference is less than 10 percent, or within the $\pm$ 20% accuracy of the cost estimate. Detailed breakdowns of the plant investments are shown in Tables 6.1.1 and 6.1.2.

The field construction costs for the individual process and offsite units are given in Tables 6.1.3 and 6.1.4. A summary comparison is as follows:

Field Construction Costs, MM\$

<u>Base Case</u>	<u>I</u>	<u>II</u>
Process Units		
Gasification et al	430	430
Gasoline, etc., Production	132	180
SNG Production	<u>26</u>	<u>26</u>
Sub-Total	588	636
Offsite Units		
Oxygen Facilities	110	110
Steam Facilities	126	148
Water Facilities	68	76
Catalyst Preparation	0	28
Other	<u>146</u>	<u>142</u>
Sub-Total	450	504
Infrastructure	<u>46</u>	<u>46</u>
Total	1,084	1,186

The above comparison emphasizes the difficulty in using the total plant investment when comparing technical changes in only one sector of the plant complex. Again, the total cost comparison shows Base Case I to have less than a 10 percent improvement over Base Case II. When comparing those investments related to the gasoline production sector only, the Mobil methanol technology, however, shows a significant 36% reduction in the investment over that required by the Fischer-Tropsch technology. Moreover, a partial offsite breakdown also shows an investment saving of about 30 percent.

6.1.3 Sensitivity Cases

The total plant investments for the sensitivity cases are estimated to be:

<u>Sensitivity Case</u>	<u>MM \$</u>
I-A	1,587
I-B	1,986
I-C	1,680
II-A	1,727

TABLE 6.1.1
PLANT INVESTMENT BREAKDOWN - BASE CASE I

	<u>Investment, MM \$</u>
<u>Depreciable Capital</u>	
Process Units	588
Offsite Units	450
Infrastructure	<u>46</u>
Sub-Total, Field Construction Cost	1,084
Engineering & Design	<u>125</u>
Sub-Total	1,209
Contractor Fee	<u>14</u>
Sub-Total, Contractor Cost	1,223
Other Project Costs	
Capitalized Spare Parts	4
Process Unit Data Logger	3
Construction Workers' Camp	73
Overtime Premium	134
Project Management	31
Builder's All-Risk Insurance	<u>3</u>
Sub-Total	1,471
Estimating Allowance	<u>220</u>
Sub-Total	1,691
Catalyst	0
Royalties	<u>17</u>
Total Depreciable Capital	1,708
<u>Expense Capital</u>	
Sales Tax	16.5
Catalyst & Chemicals	6.0
Environmental Studies & Impact Reports	<u>2.0</u>
Total Expense Capital	24.5

TABLE 6.1.2
PLANT INVESTMENT BREAKDOWN - BASE CASE II

	<u>Investment, MM \$</u>
<u>Depreciable Capital</u>	
Process Units	636
Offsite Units	504
Infrastructure	<u>46</u>
Sub-Total, Field Construction Cost	1,186
Engineering & Design	<u>136</u>
Sub-Total	1,322
Contractor Fee	<u>16</u>
Sub-Total, Contractor Cost	1,338
Other Project Costs	
Capitalized Spare Parts	4
Process Unit Data Logger	3
Construction Workers' Camp	80
Overtime Premium	147
Project Management	33
Builder's All-Risk Insurance	<u>3</u>
Sub-Total	1,608
Estimating Allowance	<u>242</u>
Sub-Total	1,850
Catalyst	0
Royalties	<u>13</u>
Total Depreciable Capital	1,863
<u>Expense Capital</u>	
Sales Tax	18.5
Catalyst & Chemicals	3.0
Environmental Studies & Impact Reports	<u>2.0</u>
	23.5

TABLE 6.1.3
UNIT INVESTMENT BREAKDOWN - BASE CASE I

<u>Unit No.</u>	<u>Description</u>	<u>Field Construction Cost, MM \$</u>
Process Units		
101	Gasification	200.7
102	Raw Gas Shift	12.8
103	Raw Gas Cooling	13.3
104	Shifted Gas Cooling	6.0
105	Gas Purification	71.7
106	Sulfur Recovery	59.0
107	Gas Liquor Separation	18.6
108	Phenol Recovery	14.0
109	Ammonia Recovery	18.8
110	Methanol Synthesis	54.4
111	H ₂ Recovery	2.6
112	Methanation	18.0
113	CO ₂ Removal	6.8
114	SNG Drying	0.5
150	Methanol Conversion	59.4
151	Naphtha Hydrotreating	3.7
152	Fractionation	3.2
153	HF Alkylation	3.5
	Miscellaneous (Control House, Pipe Rack)	21.0
	Sub-Total	588.0
Offsite Units		
121	Oxygen Production	110.1
122, 123, 125 & 126	Boiler, Main Superheater, Stackgas Precipitator and Clean-Up	122.1
124	Superheater	3.6
127	Instrument and Plant Air	0.6
128	Coal Handling	63.3
129	Ash Handling	5.6
131	BFW Preparation	16.6
132	CW Make-Up Preparation	0.5
133	CW Towers	20.0
134	Power Generation	10.3
135	Waste Water Treatment	25.4
136	Blow-Down Facilities	2.6
137	Storage	22.5
138	Interconnecting Piping	20.4
141	Refrigeration	20.8
154	Gasoline Blending	1.7
	Miscellaneous (Water Ponds, Misc. Tankage)	3.4
	Sub-Total	449.5
Infrastructure (See Paragraph 1.6.)		46.4
Total		1,083.9

TABLE 6.1.4
UNIT INVESTMENT BREAKDOWN - BASE CASE II

<u>Unit No.</u>	<u>Description</u>	<u>Field Construction Cost, MM \$</u>
Process Units		
201	Gasification, Shift, Cooling	
to	Purification & Recovery -	
209	Same as Base Case I	414.9
210	Hydrocarbon Recovery	19.4
211	H ₂ Recovery	8.8
212	Methanation	12.0
213	CO ₂ Removal	9.2
214	SNG Drying & Compression	3.1
250	F-T Synthesis	76.4
251	Naphtha Hydrotreating	3.7
252	F-T Product Fractionation	14.5
253	F-T Product Hydrogenation	6.6
254	Hydrotreated Product Fractionation	3.8
255	Catalytic Reforming	7.3
256	C ₅ /C ₆ Isomerization	5.3
257	Catalytic Polymerization	4.1
258	HF Alkylation	2.2
259	Poly Gasoline Hydrogenation	2.4
260	Light Ends Recovery	2.2
261	H ₂ Purification	2.7
262	Alcohol Recovery	11.0
	Miscellaneous (Control House, Pipe Rack)	26.6
	Sub-Total	636.2
Offsite Units		
221	Oxygen Production	110.1
222, 223, 225 & 226	Boiler, Main Superheater, Stackgas Precipitator and Clean-Up	146.6
224	Superheater	1.6
227	Instrument and Plant Air	0.6
228	Coal Handling	65.6
229	Ash Handling	5.8
231	BFW Preparation	20.6
232	CW Make-Up Preparation	0.5
233	CW Towers	23.3
234	Power Generation	10.6
235	Waste Water Treatment	26.3
236	Blow-Down Facilities	2.7
237	Storage	8.6
238	Interconnecting Piping	20.4
241	Refrigeration	27.1
270	Gasoline Blending	2.0
271	F-T Catalyst Preparation	27.7
	Miscellaneous (Water Ponds, Misc. Tankage)	3.4
	Sub-Total	503.5
Infrastructure (See Paragraph 1.6.)		46.4
Total		1,186.1

Abbreviated investment breakdowns are given in Table 6.1.5. For comparison, the base case breakdowns are also included. The effect on plant investment brought about by the technical sensitivity change is readily apparent: for example, (1) the lower gasoline production sector investment for the fluid-bed, versus fixed-bed, operation in the Mobil methanol conversion technology, (2) the increased investments required when shifting to a gasoline-only plant production mode of operation, and (3) the potential of the Mobil direct route technology to have a plant investment comparable to one for the Mobil methanol conversion technology.

New Unit investment costs (as defined in Tables 6.1.3 and 6.1.4) for the sensitivity cases are:

<u>Sensitivity Case</u>	<u>Unit</u>	<u>Plant Construction Cost, MM \$</u>
I-A	115 - Methanol Distillation	6.6
I-B	116 - Autothermal Reforming	36.5
I-C	150 - Methanol Conversion (Fluid-Bed)	20.5
	152 - Fractionation	11.3
II-A	250 - Direct Conversion	72.4
	252 - Fractionation	5.8

TABLE 6.1.5
INVESTMENTS FOR SENSITIVITY CASES

Case	Base	-----Sensitivity-----			Base	Sens.
	<u>I</u>	<u>I-A</u>	<u>I-B</u>	<u>I-C</u>	<u>II</u>	<u>II-A</u>
Depreciable Capital, MM \$						
Process Units						
Gasification et al	430	430	467	430	430	430
Gasoline Production	132	70*	206	105	180	119
SNG Production	26	26	0	26	26	26
Sub-Total	588	526	673	561	636	575
Offsite Units						
Oxygen Facilities	110	110	145	110	110	110
Steam Facilities	126	116	148	120	148	139
Water Facilities	68	61	71	65	76	73
Catalyst Preparation	0	0	0	0	28	0
Other	146	132	159	148	142	143
Sub-Total	450	419	523	443	504	465
Infrastructure	46	46	46	46	46	46
Sub-Total, Field Const.	1,084	991	1,242	1,050	1,186	1,086
Engineering & Design	139	126	160	135	152	139
Sub-Total, Contractor Cost	1,223	1,117	1,402	1,185	1,338	1,225
Other Project Costs	248	227	284	240	270	248
Estimating Allowance	220	202	252	214	242	220
Royalties	17	17	19	17	13	13
Total	1,708	1,563	1,957	1,656	1,863	1,706
Expense Capital, MM \$						
Sales Tax	16	16	19	16	19	16
Catalyst & Chemicals	6	6	8	6	3	3
Environmental Studies, etc.	2	2	2	2	2	2
Total	24	24	29	24	24	21

*Methanol production

6.2 WORKING CAPITAL

6.2.1 Bases

The working capital bases outlined in the Guidelines yield a working capital which can be classified as incremental, or is appropriate for large integrated corporations. For smaller, "single product" companies, the working capital probably should be larger.

The working capital calculation bases are:

- (a) Coal Inventory @ 30 days - Generally, in coal processing studies of this type, a coal dead storage of 30 days is assumed.
- (b) Intermediate and Product Inventory @ one-half storage capacity - There is no gas storage. SNG storage is assumed to be included in the natural gas pipeline distribution system.
- (c) Materials and Supplies @ 0.9 percent of contractor cost - A portion of the total is shifted to Capitalized Spare Parts in the Depreciable Investment.
- (d) Net Receivables @ 1/24th of annual product revenue.
- (e) Cash-on-Hand @ 3 weeks coal and operating costs - This item, omitted in the Guidelines, is generally included in the working capital estimate.

For the purposes of determining the product revenue, or product/intermediate values, the following simplified calculation on an annual basis is used:

Product Revenue = Coal Cost + Operating Cost - Byproduct

Credit + Capital Charges @ 20 percent of Plant Investment

No feedback of the actual capital charges as determined by the specific bases used for the economic calculation is employed. Intermediate product values are prorated approximately from the calculated product value.

6.2.2 Study Cases

The following working capital estimates have been developed for the study cases:

<u>Case</u>	<u>Working Capital, MM \$</u>
I	58
I-A	50
I-B	76
I-C	57
II	60
II-A	52

Breakdowns of the base case working capitals are given in Table 6.2.1.

TABLE 6.2.1
WORKING CAPITAL BREAKDOWNS FOR BASE CASES

<u>Base Case</u>	<u>I</u>	<u>II</u>
Item, MM \$		
Coal Inventory	6	6
Intermediate/Product Inventory	11	9
Warehouse Inventory	8	8
Net Receivables	22	24
Cash-on-Hand	11	12
Platinum Inventory*	<u>0</u>	<u>1</u>
	58	60

*For Reforming Unit 255 catalyst

6.3 COAL AND OPERATING COSTS

6.3.1 Coal Price

The cost of Wyoming strip sub-bituminous coal delivered to a neighboring conversion plant is assumed to be \$7.00/T, based on available price data for Gillette strip coal.

6.3.2 Operating Labor

The estimated hourly operating manpower for the base cases is shown in Table 6.3.1.

The greater complexity of units required by the Fischer-Tropsch technology results in almost a 50 percent greater operating labor requirement.

The estimates of the total hourly operating manpower for the sensitivity cases are:

<u>Sensitivity Case</u>	<u>Manpower, Men</u>
I-A	241
I-B	257
I-C	259
II-A	290 (no catalyst plant)

In Sensitivity Case II-A, the less complex product upgrading required for the Mobil direct conversion technology reduces this sector's manpower from the 62 men required in Base Case II to 7 men.

An annual wage of \$22,000 is used. It is based on the Guidelines hourly rate of \$6.70/hr (1/76), escalated to \$7.85 (10/77) at 9%/yr, and includes a burden factor of 35 percent. In addition, supervision is added at 20 percent yielding an equivalent annual rate of \$26,450.

6.3.3 Maintenance

The annual maintenance cost factors identified in the Guidelines have been adopted:

	<u>Factor, %</u>
(a) Coal preparation, gasification and ash handling	6.0
(b) Shift conversion, gas cooling, purification, byproduct recovery, methanation, SNG compression and drying, oxygen plant, stack gas treating and water treating	3.0
(c) All other	1.0

TABLE 6.3.1
OPERATING MANPOWER FOR BASE CASES

<u>Unit</u>	<u>Description</u>	<u>Manpower Shift*</u>	<u>Day</u>
BASE CASE I			
128	Coal Preparation & Storage	27.5	-
101,102,103,104 & 129	Gasification and Syn Gas Shift and Cooling	80	4
105,106 & 141	Syn Gas Purification	17.5	-
110,112,113 & 114	Methanol & SNG Synthesis	13.5	-
111,150,151,152, 153 & 154	Gasoline Synthesis	27	2
107,108,109,135 & 136	Effluent Water Treating	49.5	-
121,122, etc.	O ₂ Plant, Utilities, etc.	<u>36</u>	<u>4</u>
	Total	251	10
BASE CASE II			
228	Coal Preparation & Storage	27.5	-
201,202,203,204 & 229	Gasification and Syn Gas Shift and Cooling	80	4
205,206 & 241	Syn Gas Purification	17.5	-
250,252	F-T Synthesis	45	4
211,251,253 thru 262	Product Upgrading	62	2
210,212,213 & 214	SNG Synthesis	13.5	-
207,208,209,235 & 236	Effluent Water Treating	49.5	-
221,222, etc.	O ₂ Plant, Utilities, etc.	36	4
271	F-T Catalyst Preparation	<u>28</u>	<u>13</u>
	Total	359	27

*Shift conversion factor is 4.5 to allow for weekends, vacations, sickness, etc.

The following, additional factors have been assumed:

<u>Unit</u>	<u>Factor, %</u>
Methanol Synthesis	3.0
Autothermal Reformer	3.0
Methanol Conversion	4.0
F-T Synthesis	5.0
F-T Catalyst Preparation	6.0
Direct Synthesis	5.0
Hydrocarbon Upgrading	3.0
Coal Boiler, etc.	3.0

The composite factors for the study cases are as follows:

<u>Case</u>	<u>Factor, %</u>
I	3.5
I-A	3.5
I-B	3.5
I-C	3.5
II	3.6
II-A	3.6

The annual maintenance is calculated and composed as follows:

$$\text{Total, MM \$ / yr} = (\text{Contractor Cost, MM \$} + 15\% \text{ Estimating Allowance}) \times (\text{Maintenance Factor})$$

$$\text{Labor} = 60\%; \text{ Materials} = 40\%$$

No extras are added for supervision, burden, freight, sales tax, etc.

6.3.4 Catalyst and Chemicals

The following annual catalyst and chemical expenses have been estimated for the study cases:

<u>Case</u>	<u>Cat. & Chem. Expenses, MM \$ / yr</u>
I	5.1
I-A	4.6
I-B	6.0
I-C	5.1
II	3.8
II-A	3.4

In addition, the cost of the iron ore (magnetite) raw material for F-T Catalyst Preparation Unit 271 in Base Case II has been estimated only as the transportation cost of the ore from Minnesota and the transportation cost of spent catalyst (iron pigs) back to Chicago. No iron loss and relative ore value shift to a steel mill have been assumed. This cost amounts to 2.1 MM \$/yr.

The difference between the two base cases is principally the methanol synthesis catalyst make-up cost.

No catalyst costs have been included for the Mobil technologies.

6.3.5 Other

The bases for other operating cost elements are as follows:

- (a) Water - Guidelines \$0.40/M gal (1/76) escalated to \$0.45/M gal (10/77) at 6%/yr
- (b) Purchased Power - 1.65 ¢/KWH
- (c) Supplies - 30% of operating labor + maintenance materials
- (d) Administration and General Overhead - 60% of total labor, including supervision
- (e) Local Taxes and Insurance - 2.7% of total depreciable capital less royalty

6.3.6 Study Cases

The following coal and operating costs are calculated for the study cases:

<u>Case</u>	<u>Costs, MM \$/yr</u>	
	<u>Coal</u>	<u>Operating</u>
I	64.3	131.7
I-A	64.3	121.9
I-B	64.5	149.2
I-C	64.3	128.2
II	65.3	151.5
II-A	64.3	133.8

Items of the operating costs for the base cases are listed in Table 6.3.2.

TABLE 6.3.2
BREAKDOWNS OF OPERATING COSTS FOR BASE CASES

<u>Base Case</u>	Operating Cost, MM \$/yr	
	<u>I</u>	<u>II</u>
Item		
Labor plus Supervision		
Operating	6.9	10.2
Maintenance	<u>29.5</u> <u>36.4</u>	<u>33.2</u> <u>43.4</u>
Water	1.4	1.4
Supplies		
Operating	1.7	2.5
Maintenance	<u>19.7</u> <u>21.4</u>	<u>22.2</u> <u>24.7</u>
Catalyst & Chemicals	5.0	5.9
Administration & General Overhead	21.9	26.1
Local Taxes & Insurance	<u>45.6</u>	<u>50.0</u>
Total	131.7	151.5

6.4 BYPRODUCT CREDITS

6.4.1 Bases

The following bases for the byproduct credit calculations are used in the cost estimation:

- (a) Sulfur - Guidelines 25 \$/long T escalated to 25 \$/short T @ 6%/yr
- (b) Ammonia - Guidelines 140 \$/T escalated to 155 \$/T @ 6%/yr
- (c) Power - 0.0045 \$/KWH - Value is estimated to be only equivalent to backing out coal in an existing power plant.
- (d) Coal Fines - 5.25 \$/T @ Guidelines 75% of coal price

6.4.2 Study Cases

The following byproduct credits are calculated for the study cases:

<u>Case</u>	<u>Byproduct Credits, MM \$/yr</u>
I	8.2
I-A	8.7
I-B	5.9
I-C	8.8
II	6.0
II-A	6.5

Cases having the higher byproduct credits are those showing a surplus of coal fines, as shown in Table 6.4.1 in which the byproduct credit breakdowns for the base cases are presented.

TABLE 6.4.1
BREAKDOWNS OF BYPRODUCT CREDITS FOR BASE CASES

<u>Base Case</u>	Byproduct Credits, MM \$/yr	
	<u>I</u>	<u>II</u>
Item		
Sulfur	0.5	0.5
Ammonia	5.4	5.4
Power	0.2	0.1
Coal Fines	<u>2.1</u>	<u>0.0</u>
	8.2	6.0

SECTION 7
ECONOMIC EVALUATION

7.1 BASES

7.1.1 Scope

As recommended in the Gas Cost Guidelines, unit costs for each case are developed for both equity-and utility-type financing. For each type of financing, unit costs are calculated on both a thermal product and a multiple products basis. In the former, the unit cost is based on the total thermal (Btu) product yield. This cost assumes each Btu shares equally the plant cost. In the multiple products calculation, values for products other than gasoline are projected, and the gasoline unit cost is then calculated. Again, the principal on which the bases have been adopted is to yield realistic unit costs. Pertinent factors are analyzed in a fairly detailed sensitivity development around Base Case I.

7.1.2 Equity Financing Bases

The following bases are used in the equity financing calculation:

- (a) Plant Operating Life @ 20 years
- (b) Depreciable Life @ 13 years, ADR
- (c) Construction Timing @ 6 years - This timing assumes year 1 is required for environmental studies, year 2 is required for impact reports and plant design and engineering, and years 3 through 6 are required for plant construction. Investment payout is:

<u>Year</u>	<u>% of Investment</u>
1	0, 1 MM \$ for environmental studies
2	5, 1 MM \$ for environmental studies
3	24
4	30
5	25
6	16

- (d) Other Expenditures - Royalty in year 6
Working Capital in year 7 with recovery in last year.
Catalyst & Chemical Fill in year 7
- (e) Investment Credit @ 7% in year of expenditure
- (f) Depreciation Method - The sum-of-the-year digits procedure is used. The "half-rate in first year" convention is practiced. (Independent of when the plant qualifies for depreciation during the first year of operating the allowable depreciation taken is one-half the full year rate.)

- (g) Income Taxes - Federal @ 48%
State @ 7.5%
Total @ 51.9%
- (h) Start-Up Penalty - Only one-half production is assumed in first operating year with full expenses, except for the variable costs, such as coal, water, supplies, etc., which are expensed at one-half rate.
- (i) Unit Cost Basis @ 12% DCF

The calculation procedure is not the same as outlined in the Guidelines and the bases are significantly different.

7.1.3 Utility Financing Bases

The following bases are used in the utility financing calculation:

- (a) Plant Operating Life @ 20 years
- (b) Depreciable Life @ 20 years
- (c) Construction Timing @ 5 years - The investment payout factors are 5/24/30/25/16 as percent of the plant investment. With simple interest at 9%, the average spending period becomes 2.77.
- (d) Investment Credit - Not used
- (e) Depreciation Method - Straight Line
- (f) Income Taxes @ 51.9%
- (g) Start-Up Penalty @ 20% of the sum of the annual coal and operating costs
- (h) Debt/Equity Ratio @ 75/25
- (i) Unit Cost Basis @ 9% interest on debt capital and 15% return after taxes on equity capital

The calculating procedure is the same as outlined in the Guidelines and only the income tax and average spending period bases differ than those recommended.

7.1.4 Product Pricing Bases

The following product pricing bases have been adopted when calculating the economics in the multiple products mode:

- (a) SNG: Since over 50 percent of the thermal product is SNG in the coproduction cases, the selection of the plant gate price for SNG is quite critical. The basis chosen is the cost of SNG from an equivalent Wyoming coal, Lurgi gasification plant. To arrive at the prices cited below, a sensitivity case was developed as described in Appendix D.

Equity Basis	-	6.17 \$/MM Btu
Utility Basis	-	4.51 \$/MM Btu

- (b) C₃ LPG: Since this product is expected to be marketed for heating, the plant gate SNG prices have been assigned.

Equity Basis		6.17 \$/MM Btu
Utility Basis	-	4.51 \$/MM Btu

- (c) Butanes: The butanes can be used for fuel, gasoline blending or alkylation feed. It has been assumed that their end use will be gasoline. Historically, butanes for this use have a lower refinery plant gate value than posted for 10 RVP gasoline, e.g., \$0.25/Bbl. If this delta is used with the high cost gasolines in this study, however, the butanes will then have a thermal value higher than calculated for the gasoline. This situation is unreasonable. Consequently, an arbitrary 30 ¢/MM thermal value delta below gasoline has been assumed.

- (d) Diesel and Fuel Oils: For these Fischer-Tropsch products, typical refinery gate differences between motor gasoline and heavier products have been assumed.

		<u>Delta, \$/Bbl</u>
Diesel Oil	-	1.70
Fuel Oil	-	3.50

- (e) Alcohols: The estimated plant gate value for the Fischer-Tropsch refined alcohol mixture is 15 ¢/lb without any market restrictions. (With market restrictions, this value could drop to about 13 ¢/lb.) The basis is projected Gulf Coast prices with transportation from Wyoming backed out.

The above bases should be considered only as one of the possible scenarios for the market place from 10 to 15 years in the future. They do indicate, however, the effect on the gasoline unit cost when values or prices are superimposed on the coproducts.

Because of the uncertainty in extrapolating historical refinery price deltas at the current price levels for product quality differences to deltas for the high cost coal derived products, this consideration has been omitted. For example, credit should be recognized for the above target octanes of the gasolines produced by the Mobil technologies and, perhaps, for the excellent quality F-T diesel and fuel oils.

7.2 INTER-STUDY COMPARISONS

In the development of the study bases, considerable effort has been made not only to identify technical and economic differences between the Fischer-Tropsch and Mobil methanol conversion technologies, but also to obtain with available technology a realistic gasoline cost in current dollars. Except for the Mobil technology, scale-up from laboratory or bench-scale data for second generation technology and "conceptual" engineering have not been incorporated into the plant designs. Moreover, items such as construction time, start-up difficulty, construction camp, construction labor premium and Wyoming environment have been introduced into the economic bases. Consequently, direct inter-study comparisons are, at best, extremely difficult.

We specifically caution making economic comparisons between the results in this report and other results reported in the literature. Literature reported studies frequently have been developed to provide rough scoping type estimates useful for planning and technology sales. The cost data are generally rough estimates, are often computer based, reflect unproven "conceptual" engineering without an engineering contingency factor and do not reflect realistic market place costs. With these bases costs are usually very optimistic, especially if the economic bases follow the Gas Cost Guidelines. The costs in this study have been developed together by a contractor and an operating company, and hence, we believe they do represent realistic bottom line data. If comparisons to other technologies are needed, it is suggested that the cases be developed on a consistent bases, preferably by the same analyst.

To aid in inter-study comparisons, Base Case I economics have been reworked following the Gas Cost Guidelines. These data are included in Sub-Section 7.4.

7.3 BASE CASES

The unit costs calculated for the base cases are reported in Table 7.3.1.

Economics favor the Mobil methanol conversion technology from 3 to 15 \$/Bbl of gasoline, depending upon the cost calculating and product pricing bases. Placing the assumed price restrictions on the coproducts gives the new technology a substantial advantage over the less selective Fischer-Tropsch technology.

The breakdowns of the thermal unit costs clearly show the capital intensive nature of coal processing; from 60 to 65 percent of the unit cost can be attributed to capital charges. The relatively inexpensive strip coal results in a coal cost of only about 10 percent. Eventhough the annual operating expenses are quite substantial, they only amount to 20-30 percent of the unit cost.

Cash flows representing the 12% DCF cost calculation for the base cases are given in Tables 7.3.2 and 7.3.3.

TABLE 7.3.1
UNIT COSTS FOR BASE CASES

<u>Base Case</u>	<u>I</u>	<u>II</u>
Equity Bases @ 12% DCF		
Thermal Product, \$/MM Btu		
Capital Charges	4.59	5.07
Start-Up Penalty	0.35	0.38
Coal Cost	0.70	0.72
Operating Expenses	1.44	1.68
Byproduct Credit	<u>(0.09)</u>	<u>(0.07)</u>
Total	6.99	7.78
Equivalent Gasoline Unit Cost, ¢/gal	85	93
Multiple Products		
Gasoline Unit Cost, ¢/gal	98	133
Utility Basis		
Thermal Product, \$/MM Btu		
Capital Charges	2.98	3.28
Start-Up Penalty	0.05	0.06
Coal Cost	0.70	0.72
Operating Expenses	1.44	1.68
Byproduct Credit	<u>(0.09)</u>	<u>(0.07)</u>
Total	5.08	5.67
Equivalent Gasoline Unit Cost, ¢/gal	62	68
Multiple Products		
Gasoline Unit Cost, ¢/gal	71	94

TABLE 7.3.2
EQUITY CASH FLOW FOR BASE CASE I
(12% DCF)

DCFCASH DETAIL (\$/MM Btu = 6.99)

YEAR	GROSS REVENUE MM \$	DEPR. INVEST MM \$	OP.EXP. + FEED MM \$	INCOME TAX MM \$	SALE TX +SEV.TX MM \$	WORKING CAPITAL MM \$	NETCASH INFLOW MM \$	DEPRE- CIATION MM \$
1	.000	.000	1.000	-.519	.000	.000	-.481	.000
2	.000	84.550	1.000	-6.866	.825	.000	-79.509	.000
3	.000	405.840	.000	-30.463	3.959	.000	-379.335	.000
4	.000	507.300	.000	-38.079	4.948	.000	-474.169	.000
5	.000	422.750	.000	-31.733	4.124	.000	-395.141	.000
6	.000	287.560	.000	-20.309	2.639	.000	-269.890	.000
7	324.871	.000	160.830	21.819	.000	58.000	84.222	122.000
8	649.742	.000	195.960	113.748	.000	.000	340.035	234.615
9	649.742	.000	195.960	123.489	.000	.000	330.293	215.846
10	649.742	.000	195.960	133.230	.000	.000	320.552	197.077
11	649.742	.000	195.960	142.971	.000	.000	310.811	178.308
12	649.742	.000	195.960	152.713	.000	.000	301.070	159.538
13	649.742	.000	195.960	162.454	.000	.000	291.328	140.769
14	649.742	.000	195.960	172.195	.000	.000	281.587	122.000
15	649.742	.000	195.960	181.936	.000	.000	271.846	103.231
16	649.742	.000	195.960	191.677	.000	.000	262.105	84.462
17	649.742	.000	195.960	201.419	.000	.000	252.364	65.692
18	649.742	.000	195.960	211.160	.000	.000	242.622	46.923
19	649.742	.000	195.960	220.901	.000	.000	232.881	28.154
20	649.742	.000	195.960	230.642	.000	.000	223.140	9.385
21	649.742	.000	195.960	235.513	.000	.000	218.269	.000
22	649.742	.000	195.960	235.513	.000	.000	218.269	.000
23	649.742	.000	195.960	235.513	.000	.000	218.269	.000
24	649.742	.000	195.960	235.513	.000	.000	218.269	.000
25	649.742	.000	195.960	235.513	.000	.000	218.269	.000
26	649.742	.000	195.960	235.513	.000	-58.000	276.269	.000
TOT	12669.973	1708.000	3886.070	3545.463	16.495	.000	3513.945	1708.000

TABLE 7.3.3
EQUITY CASH FLOW FOR BASE CASE II
(12% DCF)

DCFCASH DETAIL (\$/MM Btu = 7.78)

YEAR	GROSS REVENUE MM \$	DEPR. INVEST MM \$	OP. EXP. + FEED MM \$	INCOME TAX MM \$	SALE TX + SEV. TX MM \$	WORKING CAPITAL MM \$	NETCASH INFLOW MM \$	DEPRE- CIATION MM \$
1	.000	.000	1.000	- .519	.000	.000	- .481	.000
2	.000	92.500	1.000	- 7.474	.924	.000	- 86.950	.000
3	.000	444.000	.000	- 33.382	4.435	.000	- 415.053	.000
4	.000	555.000	.000	- 41.727	5.544	.000	- 518.817	.000
5	.000	462.500	.000	- 34.773	4.620	.000	- 432.347	.000
6	.000	309.000	.000	- 22.255	2.957	.000	- 289.702	.000
7	355.278	.000	176.790	23.571	.000	60.000	94.917	133.071
8	710.635	.000	216.810	123.479	.000	.000	370.345	255.907
9	710.635	.000	216.810	134.105	.000	.000	359.720	235.434
10	710.635	.000	216.810	144.730	.000	.000	349.095	214.962
11	710.635	.000	216.810	155.355	.000	.000	338.469	194.489
12	710.635	.000	216.810	165.980	.000	.000	327.844	174.016
13	710.635	.000	216.810	176.606	.000	.000	317.219	153.544
14	710.635	.000	216.810	187.231	.000	.000	306.594	133.071
15	710.635	.000	216.810	197.856	.000	.000	295.968	112.599
16	710.635	.000	216.810	208.481	.000	.000	285.343	92.126
17	710.635	.000	216.810	219.107	.000	.000	274.718	71.654
18	710.635	.000	216.810	229.732	.000	.000	264.093	51.181
19	710.635	.000	216.810	240.357	.000	.000	253.467	30.709
20	710.635	.000	216.810	250.982	.000	.000	242.842	10.236
21	710.635	.000	216.810	256.295	.000	.000	237.530	.000
22	710.635	.000	216.810	256.295	.000	.000	237.530	.000
23	710.635	.000	216.810	256.295	.000	.000	237.530	.000
24	710.635	.000	216.810	256.295	.000	.000	237.530	.000
25	710.635	.000	216.810	256.295	.000	.000	237.530	.000
26	710.635	.000	216.810	256.295	.000	60.000	297.530	.000
TOT	13857.336	1863.000	4298.180	3855.214	18.480	.000	3822.462	1863.000

7.4 INTER-STUDY COMPARISON SUB-CASE

As a guide in making inter-study economic comparisons, a sub-case has been developed from Base Case I data following exactly, except for two items, the economic bases and calculations outlined in the Gas Cost Guidelines. The two exceptions are the usage of a 7 \$/T price for Wyoming strip coal and October, 1977 costs. The results are briefly described below:

Total Investment, MM \$ 1,295

Equity Basis @ 12% DCF

Thermal Product

Equivalent Gasoline, ¢/gal 69 (5.65 \$/MM Btu)

Multiple Product

Gasoline, ¢/gal 78 (SNG @ 4.98 \$/MM Btu)

Utility Basis

Thermal Product

Equivalent Gasoline, ¢/gal 49 (4.03 \$/MM Btu)

Multiple Product

Gasoline, ¢/gal 56 (SNG @ 3.58 \$/MM Btu)

Investment items omitted include overtime premium and project management. A construction camp cost, however, has been included. Working capital is reduced. Construction costs are based on Gulf Coast prices and productivity. Construction worker base wage rate is 10.54 \$/hr. Timing does not reflect the requirement for environmental studies, etc. Economic bases include income tax at 48%, no investment credit and 16 year depreciable life. The stream factor is reduced to 90 percent. These bases result in about a 20 percent reduction in the unit costs derived from the bases adopted for this study.

We are not suggesting that the above gasoline costs be used without reservation in inter-study comparisons, but offer them to show that the basic study data can, indeed, lead to relatively low gasoline costs. The study assumptions to arrive at realistic costs for a Wyoming location plant have a severe effect on the venture economics.

7.5 BASE CASE I ECONOMIC SENSITIVITIES

Since Base Case I has the better economics, economic sensitivities have been developed only for this case. Because of the similar nature between the study cases, the Base Case I effects described below would also be generally applicable to all cases.

7.5.1 Economic Bases

Table 7.5.1 is a listing of the unit cost sensitivity towards the various economic bases under equity financing. The effect of the DCF rate of return is shown in Figure 7.5.1. Utility financing sensitivities are given in Table 7.5.2. Because in most sensitivities, e.g., operating life, coal price, debt/equity ratio, etc., it seems meaningless to calculate a gasoline cost at the sensitivity basis while using an SNG cost at the standard basis, sensitivities for the multiple products calculation have not been made.

Most important in the multiple products calculation is the SNG market price assumed. The sensitivity of the gasoline cost to this item is presented in Figure 7.5.2. It is applicable to both financing methods.

7.5.2 Construction Timing

We estimate that approximately 36 million manhours would be required to erect each plant complex. If the construction period assumed is four years (study basis), then at a 40 hour week, about 5,200 men would be required. With a 14 hour overtime premium (study basis) about 4,000 men would be required. These numbers represent an average; actually, the manpower rises and falls with the job requirements, peaking, generally at about 20 to 30% above the average. In a remote area as Wyoming, a more pessimistic assumption for manpower availability might be an average of only 2,000 men, thereby increasing the construction timing to 8 years.

The effect of construction timing on economics is shown in Table 7.5.3.

The stretchout of plant construction from 4 to 8 years has a calamitous effect on the gasoline cost, even though the deltas shown do not reflect any escalation. If this factor were introduced, they would be even more dramatic. Also, the time-value effect in the DCF calculation results in a more severe, but realistic, effect for a construction stretchout.

TABLE 7.5.1
SENSITIVITIES FOR BASE CASE I
EQUITY FINANCING
(12% DCF)

<u>Calculation Basis</u>	-----Unit Cost Delta-----	
	<u>Thermal Product, \$/MM Btu</u>	<u>Equivalent Gasoline, ¢/gal*</u>
25 yr. Operating Life (+5 yr.)	-0.24	-3
No Investment Credit	+0.47	+5½
10 year Depreciable Life	-0.15	-2
16 year Depreciable Life	+0.14	+1½
35% FIT (-13%)	-0.66	-8
+ 20% Depreciable Investment (+ 338 MM \$)	+0.94	+11½
Triple Working Capital (+ 116 MM \$)	+0.30	+3½
+ 2 \$/T Coal Price	+0.20	+2
+ 25% Operating Cost (+ 33.0 MM \$/yr.)	+ 0.36	+4½

*At 5.1 MM Btu/Bbl

TABLE 7.5.2
SENSITIVITIES FOR BASE CASE I
UTILITY FINANCING

<u>Calculation Basis</u>	-----Unit Cost Delta-----	
	<u>Thermal Product, \$/MM Btu</u>	<u>Equivalent Gasoline, ¢/gal*</u>
12% Interest on Debt (+3%)	+0.46	+5½
20% Return on Equity (+5%)	+0.33	+4
50/50 Debt/Equity Ratio	+0.70	+8½
60/40 Debt/Equity Ratio	+0.42	+5
35% FIT (-13%)	-0.25	-3
+ 20% Depreciable Investment (+ 338 MM \$)	+0.56	+7
Triple Working Capital (+116 MM \$)	+0.18	+2
+ 2 \$/T Coal Price	+0.20	+2
+ 25% Operating Cost (+ 33.0 MM \$/yr.)	+0.30	+4½

*At 5.1 MM Btu/Bbl

FIGURE 7.5.1
EFFECT OF DCF RATE OF RETURN ON UNIT COSTS
(BASE CASE 1)

FIGURE 7.5.2
EFFECT OF SNG PRICE ON GASOLINE UNIT COST
(BASE CASE 1)

Two questions, therefore, arise: (1) Can 4,000 men (study basis) ever be enticed into Wyoming? (2) Are the 14 hour overtime premium and construction camp monies sufficient inducements? We believe that, without further industrialization in the Montana, North Dakota and Wyoming area, these large scale coal processing plants might find it difficult to attract such large work forces. Moreover, an additional bonus will probably be required to ensure that the workers stay for a reasonable period. A definitive construction environment study would be made before any project funding decision.

7.5.3 Start-Up Penalty

The effect of the assumed start-up penalty on the unit cost is shown below:

Equity Financing @ 12% DCF

<u>Calculation Basis</u>	-----Unit Cost Delta-----	
	<u>Thermal Product, \$/MM Btu</u>	<u>Equivalent Gasoline, ¢/gal</u>
No Penalty	Base	Base
Gas Cost Guidelines Basis	+0.06	+ $\frac{1}{2}$
Study Basis	+0.35	+4

We believe it is unrealistic to assume that during the last year of construction, the complete plant operating personnel will be trained and all the initial operating and maintenance problems will be solved to achieve 100 percent product manufacture in the first year of operation. Past experience for first syn fuels plants, i.e., Great Canadian Oil Sands and Sasol I, indicates start-ups are slow and difficult. Although first plant learning should reduce start-up problems, the second tar sands project, Syncrude of Canada, currently under construction, has been projecting a gradual build-up in production capacity through the early operating years.

7.5.4 Use of Lurgi Gasifier Tar and Oil Products

The study bases call for the combustion of the Lurgi gasifier oil and tar in the boiler and process steam superheater. A sensitivity case has been roughly developed for backing out these materials from the boiler in Base Case I. (The separate process steam superheater, Unit 124, remains oil fired.) Additional hydrogen is removed from the methanol synthesis unit purge gas in H₂ Recovery Unit 111 for hydrotreating.

TABLE 7.5.3
CONSTRUCTION TIMING SENSITIVITY FOR BASE CASE I

<u>Calculation Basis</u>	-----Unit Cost Delta-----			
	-----Equity Basis-----		-----Utility Basis-----	
	Thermal Product, \$/MM Btu	Equivalent Gasoline, ¢/gal ⁽²⁾	Thermal Product, \$/MM Btu	Equivalent Gasoline, ¢/gal ⁽²⁾
1 yr. Construction	-1.18	-14½	-0.36	-4½
8 yr. Construction: Timing, % at 0/4/15/20/16/12/11/10/7/5 ⁽¹⁾ (all purchasing in early years)	+2.41	+29	+0.55	+6½
8 yr. Construction: Timing, % at 0/4/12/14/16/14/13/11/10/6 ⁽¹⁾	+2.06	+25	+0.47	+5½

(1) Year 1 is for environmental studies; year 2 is for engineering and design.

(2) At 5.1 MM Btu/Bbl.

The hydrotreating requirements were based on in-house information for upgrading an H-Coal 400⁺ material from 7.4 wt. % to 10.0 wt. % hydrogen into a marketable fuel oil. Approximately 5,700 B/SD of fuel oil is estimated to be produced. The overall plant thermal efficiency is increased by about 2 percent to 64 percent.

Relative to Base Case I, the coal, operating and capital costs increase about 2%, but the thermal product yield increases by 11%. The result is a 50 to 60 ¢/MM Btu reduction in the unit cost (equity basis), or about 7 ¢/gal in the equivalent gasoline cost. Consequently, in future studies, it appears economically advantageous to expand the hydrogen recovery unit and increase the hydrotreating capacity for the inclusion of the tar and oil as well as the naphtha.

In downstream processing situations where there is a purge gas stream whose quantity is determined by the nitrogen (inerts) present, this purge gas is often used as boiler fuel. We believe that a more expensive 99+ oxygen purity plant, thereby reducing the purge gas stream, would prove to be economical.

7.6 SENSITIVITY CASES

The unit costs calculated for the sensitivity cases are reported in Table 7.6.1.

7.6.1 Sensitivity Case I-A

Not surprising, the thermal unit cost for methanol production is about 10% lower than the cost for gasoline (via fluid-bed operation). The fuel methanol cost ranges from 29 to 40 ¢/gal, depending upon the financing basis. Since the values placed on SNG are identical to the thermal product unit costs for this case, there is no cost increase when the calculation is on the multiple products basis.

7.6.2 Sensitivity Case I-B

If SNG is a valuable and marketable product, converting the Lurgi gasifier methane into additional synthesis gas is not an economical scheme to increase gasoline production. The gasoline cost increases to a range of 85 to 125 ¢/gal, depending primarily on the financing basis. The cost spread for the multiple products calculation is small because of the relatively insignificant C₃ and C₄ yields. The SNG breakeven value for Base Case I below which methane reforming does become economical is about 4.45 \$/MM Btu (equity) and 3.25 \$/MM Btu (utility).

7.6.3 Sensitivity Case I-C

Fluid-bed operation as projected for this study has an economic advantage, up to 5 ¢/gal, over the fixed-bed unit. Furthermore, if the fixed-bed butane yield in excess of the requirements for gasoline pressuring can be marketed only as an LPG, the fluid-bed advantage increases by an additional 1½ ¢/gal.

7.6.4 Sensitivity Case II-A

The projected operation and yields indicate that the Mobil direct route technology has the potential of being as economical on a thermal yield basis as the Mobil methanol conversion technology. Hence, it is a significant improvement to the commercial F-T technology. Its poorer gasoline selectivity relative to the methanol route - though superior to F-T -, however, leads to higher gasoline costs on a multiple products basis; 75 to 110 ¢/gal vs. 65 to 100 ¢/gal.

Laboratory studies indicate that direct conversion can occur at a H₂/CO ratio of unity, as opposed to the 2/1 ratio required for the methanol and Fischer-Tropsch syntheses. Consequently, this technology has the possibility of being an excellent partner with the advanced, high capacity gasifiers which generally yield low H₂/CO ratios.

TABLE 7.6.1
UNIT COSTS FOR SENSITIVITY CASES

Case Products	I (Base) SNG & Gasoline	I-A SNG & Methanol	I-B Gasoline Only	I-C ⁽³⁾ SNG & Gasoline	II (Base) SNG & Gasoline	II-A SNG & Gasoline
Equity Basis @ 12% DCF						
Thermal Product						
\$/MM Btu Equivalent Gasoline, ¢/gal ⁽¹⁾	6.99 85	6.15 40(2)	9.98 121	6.80 83	7.78 93	6.99 85
Multiple Products						
Gasoline, ¢/gal	98	40(2)	124	93	133	106
Utility Basis						
Thermal Product						
\$/MM Btu Equivalent Gasoline, ¢/gal ⁽¹⁾	5.08 62	4.47 29(2)	7.24 88	4.94 60	5.67 68	5.09 62
Multiple Products						
Gasoline, ¢/gal	71	29(2)	90	67	94	76

(1) At 5.1 MM Btu/Bbl (Cases I, I-B, I-C & II-A); 5.0 MM Btu/Bbl (Case II)

(2) Methanol

(3) Fluid-Bed Operation

SECTION 8 CONCLUSIONS

1. In all areas, the Mobil methanol conversion technology is superior to the commercial Fischer-Tropsch technology for the manufacture of gasoline; i.e., it has an advantage in processing complexity, thermal efficiency, product selectivity, investment and gasoline quality. The gasoline unit cost advantage ranges from 6 to 40 ¢/gal, depending upon the pricing and financing assumptions.
2. With SNG coproduction, the October, 1977 gasoline cost for the Mobil methanol conversion technology is estimated to range between 60 to 100 ¢/gal, depending upon the pricing and financing assumptions, for the adopted study bases. Optimistic bases, such as recommended in the "Coal Gasification Commercial Concepts Gas Cost Guidelines", however, yield a gasoline cost range of 50 to 80 ¢/gal, or about 20% lower.
3. If SNG is a valuable product, it is not economical to convert the Lurgi gasifier methane into additional gasoline. For the gasoline-only operation, the gasoline cost increases by 15 to 25 ¢/gal.
4. The projected fluid-bed operation of the Mobil methanol conversion technology has an economic advantage over the fixed-bed operation. This advantage ranges from 2 to 7 ¢/gal, depending principally on the values assigned to SNG, C₃ LPG, and excess butanes. Fluid-bed operation has an advantage in engineering complexity, thermal efficiency, gasoline selectivity, gasoline octane and investment.
5. The Mobil direct route technology has the potential of being, at least, as economical as the methanol conversion technology. Presently, however, its poorer gasoline selectivity results in a higher gasoline unit cost of about 10 ¢/gal when the coproducts are valued below gasoline.
6. For large plants, it is economically advantageous to process the gasifier liquids into finished products, instead of burning them in the coal-fired boiler. The incremental cost required for upgrading is relatively small compared to the improvement in the thermal yield.

SECTION 9
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APPENDIX A
MATERIAL BALANCE - BASE CASE I

MATERIAL BALANCE

BASE CASE I

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STREAM N°		1	2.1	2	3	4
		Mined Coal	Fine coal to be sold	Fine coal to Unit 122	Well sized coal	Oxygen
	CO ₂ 44.010 lbmol/hr					
	H ₂ S 34.076 lbmol/hr					
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERTS 28.013 lbmol/hr					277.9
	O ₂ 32.000 lbmol/hr					14 060.0
	TOTAL DRY GAS lbmol/hr					14 337.9
	H ₂ O 18.016 lbmol/hr					
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS lbmol/hr					14 337.9
	TOTAL DRY GAS lb/hr					457 697
	H ₂ O lb/hr	637 815	28 700	76 725	532 390	
	METHANOL lb/hr					
	TAR lb/hr					
	OIL lb/hr					
	NAPHTHA lb/hr					
	CRUDE PHENOLS lb/hr					
	ANHYDR. AMMONIA lb/hr					
	SULPHUR lb/hr					
	OTHERS lb/hr					
	MINERALS lb/hr	116 172	5 228	13 973	96 971	
	COAL DAF lb/hr	1 523 837	68 572	183 382	1 271 963	
	SULPHUR ORGANIC lb/hr					
	NITROGEN ORGANIC lb/hr					
	FATTY ACIDS lb/hr					
	ACETONE lb/hr					
	C ₆ + lb/hr					
	TOTAL lb/hr	2 277 824	102 500	274 000	1 901 324	457 697

MATERIAL BALANCE

BASE CASE I

STREAM №		5	5.1	6	6.1	6.2
		MP Steam	BFW	Ash to disposal	Ash from Unit 129	Ash from Unit 122
	CO ₂	44.010 lbmol/hr				
	H ₂ S	34.076 lbmol/hr				
	COS	60.070 lbmol/hr				
	C ₂ H ₄	28.054 lbmol/hr				
	CO	28.010 lbmol/hr				
	H ₂	2.016 lbmol/hr				
	CH ₄	16.043 lbmol/hr				
	C ₂ H ₆	30.070 lbmol/hr				
	N ₂ +INERTS	28.013 lbmol/hr				
	O ₂	32.000 lbmol/hr				
	TOTAL DRY GAS	lbmol/hr				
	H ₂ O	18.016 lbmol/hr	94 379.2	10 591.2	1 110.12	1 110.12
	METHANOL	32.042 lbmol/hr				
	TOTAL WET GAS	lbmol/hr	94 379.2	10 591.2		
	TOTAL DRY GAS	lb/hr				
	H ₂ O	lb/hr	1 700 335	190 811	20 000	20 000
	METHANOL	lb/hr				
	TAR	lb/hr				
	OIL	lb/hr				
	NAPHTHA	lb/hr				
	CRUDE PHENOLS	lb/hr				
	ANHYDR. AMMONIA	lb/hr				
	SULPHUR (without SO ₂)	lb/hr		79	66	13
	OTHERS	lb/hr				
	MINERALS	lb/hr		110 944	96 971	13 973
	COAL	lb/hr		7 038	7 038	
	SULPHUR ORGANIC	lb/hr				
	NITROGEN ORGANIC	lb/hr				
	FATTY ACIDS	lb/hr				
	ACETONE	lb/hr				
	C ₆ +	lb/hr				
	SO ₂ (1)			2 026		
	TOTAL	lb/hr	1 700 335	190 811	140 087	124 075
	1) Mineralized but balanced as SO ₂					

MATERIAL BALANCE

BASE CASE I

STREAM №			7	8	9	10	11
			Dusty gas Liquor	Unit 103 gas liquor	Unit 104 gas liquor	Unit 105 gas liquor	Unit 107 feed
	CO ₂	44.010 lbmol/hr	1.73				1.73
	H ₂ S	34.076 lbmol/hr	0.88				0.88
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr					
	CH ₄	16.043 lbmol/hr					
	C ₂ H ₆	30.070 lbmol/hr					
	N ₂ +INERTS	28.013 lbmol/hr					
	O ₂	32.000 lbmol/hr					
	TOTAL DRY GAS	lbmol/hr	2.61				2.61
	H ₂ O	18.016 lbmol/hr	54 541.8	18 895.0	12 041.4	134.3	85 612.5
	METHANOL	32.042 lbmol/hr					
	TOTAL WET GAS	lbmol/hr					
	TOTAL DRY GAS	lb/hr	106				106
	H ₂ O	lb/hr	982 625	340 413	216 938	2 420	1 542 396
	METHANOL	lb/hr					
	TAR	lb/hr	31 951	3 604	836		36 391
	OIL	lb/hr	13 444	31 221	7 246		51 911
	NAPHTHA	lb/hr					
	CRUDE PHENOLS	lb/hr	9 959	4 248	986		15 193
	ANHYDR. AMMONIA	lb/hr	1 734	6 556	1 522		9 812
	SULPHUR	lb/hr					
	OTHERS	lb/hr					
	MINERALS	lb/hr					
	COAL DUST	lb/hr	4 113				4 113
	SULPHUR ORGANIC	lb/hr					
	NITROGEN ORGANIC	lb/hr					
	FATTY ACIDS	lb/hr	1 770	835	194		2 799
	ACETONE	lb/hr					
	C ₆ +	lb/hr					
	TOTAL	lb/hr	1 045 702	386 877	227 722	2 420	1 662 721

MATERIAL BALANCE

BASE CASE I

STREAM №	12	13	14	15	16
	Tar to Boiler	Oil from Unit 107	Unit 108 Feed	Phenols to Boiler	Unit 109 Feed
CO ₂	44.010 lbmol/hr		1.73		1.73
H ₂ S	34.076 lbmol/hr		0.88		0.88
COS	60.070 lbmol/hr				
C ₂ H ₄	28.054 lbmol/hr				
CO	28.010 lbmol/hr				
H ₂	2.016 lbmol/hr				
CH ₄	16.043 lbmol/hr				
C ₂ H ₆	30.070 lbmol/hr				
N ₂ +INERTS	28.013 lbmol/hr				
O ₂	32.000 lbmol/hr				
TOTAL DRY GAS	18.016 lbmol/hr		2.61		2.61
H ₂ O	18.016 lbmol/hr		85.482.29		85.801.34
METHANOL	32.042 lbmol/hr				
TOTAL WET GAS	1bmol/hr				
TOTAL DRY GAS	1bmol/hr		106		106
H ₂ O	1bmol/hr	2.347	1.540.049	452	1.545.797
METHANOL	1bmol/hr				
TAR	1bmol/hr	36.391			
OIL	1bmol/hr	51.711	200	200	
NAPHTHA	1bmol/hr				
CRUDE PHENOLS	1bmol/hr		15.193	14.606	587
ANHYDR. AMMONIA	1bmol/hr				9.812
SULPHUR	1bmol/hr				
OTHERS	1bmol/hr				
MINERALS	1bmol/hr				
COAL DUST	1bmol/hr	4.113			
SULPHUR ORGANIC	1bmol/hr				
NITROGEN ORGANIC	1bmol/hr				
FATTY ACIDS	1bmol/hr		2.799		2.799
ACETONE	1bmol/hr				
C ₆ +	1bmol/hr				
TOTAL	1bmol/hr	42.851	1.568.159	15.258	1.559.101

MATERIAL BALANCE

BASE CASE I

STREAM №			17	18	18.1	18.2	18.3
			NH ₃ Product	Water to Unit 135	Water from Unit 109	Water from Unit 105	Unit 105 Make up water
	CO ₂	44.010 lbmol/hr		1.73	1.73		
	H ₂ S	34.076 lbmol/hr		0.88	0.88		
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr					
	CH ₄	16.043 lbmol/hr					
	C ₂ H ₆	30.070 lbmol/hr					
	N ₂ +INERTS	28.013 lbmol/hr					
	O ₂	32.000 lbmol/hr					
	TOTAL DRY GAS	lbmol/hr		2.61	2.61		
	H ₂ O	18.016 lbmol/hr		89 535.80	85 799.12	3 736.68	3.736.68
	METHANOL	32.042 lbmol/hr					
	TOTAL WET GAS	lbmol/hr					
	TOTAL DRY GAS	lb/hr		106	106		
	H ₂ O	lb/hr	40	1 613 077	1 545 757	67 320	67 320
	METHANOL	lb/hr					
	TAR	lb/hr					
	OIL	lb/hr					
	NAPHTHA	lb/hr					
	CRUDE PHENOLS	lb/hr		587	587		
	ANHYDR. AMMONIA	lb/hr	8 521	1 291	1 291		
	SULPHUR	lb/hr					
	OTHERS	lb/hr					
	MINERALS	lb/hr					
	COAL	lb/hr					
	SULPHUR ORGANIC	lb/hr					
	NITROGEN ORGANIC	lb/hr					
	FATTY ACIDS	lb/hr		2 799	2 799		
	ACETONE	lb/hr					
	C ₆ +	lb/hr					
	TOTAL	lb/hr	8 561	1 617 860	1 550 540	67 320	67 320

MATERIAL BALANCE

BASE CASE I

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STREAM N ^o		19	20	21	22	23
		Raw gas from Unit 101	Raw gas to Unit 103	Raw gas from Unit 103	Raw gas to Unit 102	Shift gas to Unit 104
CO ₂	44.010 lbmol/hr	33 722.0	27 369.5	27 369.5	6 352.5	9 176.8
H ₂ S	34.076 lbmol/hr	157.3	127.7	127.7	29.6	29.6
COS	60.070 lbmol/hr	2.0	1.6	1.6	0.4	0.4
C ₂ H ₄	28.054 lbmol/hr	85.5	69.4	69.4	16.1	16.1
CO	28.010 lbmol/hr	21 458.5	17 416.1	17 416.1	4 042.4	1 218.1
H ₂	2.016 lbmol/hr	44 332.0	35 972.6	35 972.6	8 359.4	11 183.7
CH ₄	16.043 lbmol/hr	12 866.0	10 442.3	10 442.3	2 423.7	2 423.7
C ₂ H ₆	30.070 lbmol/hr	614.0	498.3	498.3	115.7	115.7
N ₂ +INERTS	28.013 lbmol/hr	310.3	251.8	251.8	58.5	58.5
O ₂	32.000 lbmol/hr					
TOTAL DRY GAS	lbmol/hr	113 547.6	92 149.3	92 149.3	21 398.3	24 222.6
H ₂ O	18.016 lbmol/hr	79 122.8	64 217.6	137.3	14 905.2	12 080.8
METHANOL	32.042 lbmol/hr					
TOTAL WET GAS	lbmol/hr	192 670.4	156 366.9	92 286.6	36 303.5	36 303.4
TOTAL DRY GAS	lb/hr	2 415 949	1 960 831	1 960 831	455 118	506 001
H ₂ O	lb/hr	1 425 476	1 156 945	2 474	268 531	217 648
METHANOL	lb/hr					
TAR	lb/hr	4 440	3 604		836	836
OIL	lb/hr	38 467	31 221		7 246	7 246
NAPHTHA	lb/hr	15 675	12 722	12 722	2 953	2 953
CRUDE PHENOLS	lb/hr	5 234	4 248		986	986
ANHYDR. AMMONIA	lb/hr	8 078	6 556		1 522	1 522
SULPHUR	lb/hr					
OTHERS	lb/hr					
MINERALS	lb/hr					
COAL	lb/hr					
SULPHUR ORGANIC	lb/hr	75	61	61	14	14
NITROGEN ORGANIC	lb/hr	14	11	11	3	3
FATTY ACIDS	lb/hr	1 029	835		194	194
ACETONE	lb/hr					
C ₆ +	lb/hr					
TOTAL	lb/hr	3 914 437	3 177 034	1 976 099	737 403	737 403

MATERIAL BALANCE

BASE CASE I

STREAM Nº		24	25	26	27	
		Shift gas from Unit 104	Feed to Unit 105	Naphtha to Unit 151	Acid gas to Unit 106	
CO ₂	44.010 lbmol/hr	9 176.8	36 546.3		33 646.3	
H ₂ S	34.076 lbmol/hr	29.6	157.3		157.3	
COS	60.070 lbmol/hr	0.4	2.0		2.0	
C ₂ H ₄	28.054 lbmol/hr	16.1	85.5		64.5	
CO	28.010 lbmol/hr	1 218.1	18 634.2		134.2	
H ₂	2.016 lbmol/hr	11 183.7	47 156.3		171.3	
CH ₄	16.043 lbmol/hr	2 423.7	12 866.0		406	
C ₂ H ₆	30.070 lbmol/hr	115.7	614.0		337	
N ₂ + INERTS	28.013 lbmol/hr	58.5	310.3		1.3	
O ₂	32.000 lbmol/hr					
TOTAL DRY GAS	lbmol/hr	24 222.6	116 371.9		34 919.9	
H ₂ O	18.016 lbmol/hr	39.4	176.7		42.4	
METHANOL	32.042 lbmol/hr				14.45	
TOTAL WET GAS	lbmol/hr	24 262.0	116 548.6		34 976.75	
TOTAL DRY GAS	lb/hr	506 001	2 466 832		1 508 846	
H ₂ O	lb/hr	710	3 184		764	
METHANOL	lb/hr			158	463	
TAR	lb/hr					
OIL	lb/hr					
NAPHTHA	lb/hr	2 953	15 675	15 675		
CRUDE PHENOLS	lb/hr					
ANHYDR. AMMONIA	lb/hr					
SULPHUR	lb/hr					
OTHERS	lb/hr					
MINERALS	lb/hr					
COAL	lb/hr					
SULPHUR ORGANIC	lb/hr	14	75	75		
NITROGEN ORGANIC	lb/hr	3	14	14		
FATTY ACIDS	lb/hr					
ACETONE	lb/hr					
C ₆ +	lb/hr					
TOTAL	lb/hr	509 681	2 485 780	15 922	1 510 073	

MATERIAL BALANCE

BASE CASE I

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STREAM N°			28	29	30	31	
			Unit 106 Off gas	Sulphur Prod.	Feed to Unit 110	Unit 101 Quench Water	
	CO ₂	44.010 lbmol/hr	33 646.30		2 900		
	H ₂ S	34.076 lbmol/hr	2.02				
	COS	60.070 lbmol/hr	2.00				
	C ₂ H ₄	28.054 lbmol/hr	64.50		21		
	CO	28.010 lbmol/hr	134.20		18 500		
	H ₂	2.016 lbmol/hr	174.13		46 985		
	CH ₄	16.043 lbmol/hr	410.64		12 460		
	C ₂ H ₆	30.070 lbmol/hr	337.82		277		
	N ₂ +INERT	28.013 lbmol/hr	1.30		309		
	O ₂	32.000 lbmol/hr					
	TOTAL DRY GAS (1)	lbmol/hr	34 773.99		81 452		
	H ₂ O	18.016 lbmol/hr	2 330.82			45 185.3	
	METHANOL	32.042 lbmol/hr	14.45		1.31		
	TOTAL WET GAS (1)	lbmol/hr	37 119.26		81 453.31		
	TOTAL DRY GAS	lb/hr	1 503 723		957 986		
	H ₂ O	lb/hr	41 992			814 058	
	METHANOL	lb/hr	463		42		
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr	0.18				
	iC ₄ H ₁₀	58.12 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	nC ₄ H ₁₀	58.12 lbmol/hr	0.83				
	iC ₅ H ₁₂	72.15 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	nC ₅ H ₁₂	72.15 lbmol/hr	0.07				
	C ₆ +	(2) lbmol/hr					
	NAPHTHA	lb/hr					
	SULPHUR	lb/hr		5 051			
	TOTAL	lb/hr	1 546 178	5 051	958 028	814 058	
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						

MATERIAL BALANCE

BASE CASE I

STREAM Nº			32	33	34	35	36
			Incineration gas to Unit 122	H ₂ to Unit 151	H ₂ S from Unit 151	Expansion gas from Unit 110	Unit 105 sol- vent make up
CO ₂	44.010 lbmol/hr		34 238.72			240.00	
H ₂ S	34.076 lbmol/hr		2.03		2.25		
COS	60.070 lbmol/hr		2.00				
C ₂ H ₄	28.054 lbmol/hr		64.50			1.00	
CO	28.010 lbmol/hr		134.20			28.80	
H ₂	2.016 lbmol/hr		201.21	96.22	2.83	75.76	
CH ₄	16.043 lbmol/hr		432.92	1.37	4.64	484.00	
C ₂ H ₆	30.070 lbmol/hr		338.14		0.82	9.00	
N ₂ +INERT	28.013 lbmol/hr		571.14			4.97	
O ₂	32.000 lbmol/hr		21.75				
TOTAL DRY GAS (1)	lbmol/hr		36 007.96	97.59	11.62	843.53	
H ₂ O	18.016 lbmol/hr		2 501.45			2.00	1.6
METHANOL	32.042 lbmol/hr		14.45			155.00	20.7
TOTAL WET GAS (1)	lbmol/hr		38 523.86			1 000.53	
TOTAL DRY GAS	lb/hr		1 547 356	216	242	19 725	
H ₂ O	lb/hr		45 066			36	29
METHANOL	lb/hr		463			4 966	663
C ₃ H ₆	42.08 lbmol/hr						
C ₃ H ₈	44.09 lbmol/hr		0.20		0.18		
iC ₄ H ₁₀	58.12 lbmol/hr		0.01				
1C ₄ H ₈	56.10 lbmol/hr		0.01				
nC ₄ H ₁₀	58.12 lbmol/hr		0.92		0.83		
1C ₅ H ₁₂	72.15 lbmol/hr				0.07		
1C ₅ H ₁₀	70.13 lbmol/hr						
nC ₅ H ₁₂	72.15 lbmol/hr		0.21				
C ₆ +	(2) lbmol/hr						
NAPHTHA			lb/hr				
TOTAL			lb/hr	1 592 422	216	242	24 727
(1) C ₃ + INCLUDED							692
(2) MW OF C ₆ +							

MATERIAL BALANCE

BASE CASE I

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STREAM №			37	38	39	40	41
			Feed to Unit 150	Purge gas from Unit 110	Feed to Unit 112	Feed to Unit 111	Off gas from Unit 111
	CO ₂	44.010 lbmol/hr	32.00	1 243.00	1 483.00	23.99	23.99
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr		20.00	21.00	0.39	0.39
	CO	28.010 lbmol/hr	0.20	1 240.00	1 268.80	23.93	23.93
	H ₂	2.016 lbmol/hr	0.24	8 309.00	8 288.54	160.34	64.12
	CH ₄	16.043 lbmol/hr	8.00	11 968.00	12 450.63	230.99	229.62
	C ₂ H ₆	30.070 lbmol/hr	2.00	266.00	275.00	5.13	5.13
	N ₂ +INERTS	28.013 lbmol/hr	0.03	304.00	308.97	5.87	5.87
	O ₂	32.000 lbmol/hr					
	TOTAL DRY GAS	lbmol/hr	42.47	23 350.00	24 095.94	450.64	353.05
	H ₂ O	18.016 lbmol/hr	1 382.40	1.00	3.00	0.02	0.02
	METHANOL	32.042 lbmol/hr	18 360.30	78.00	233.00	1.50	1.50
	TOTAL WET GAS	lbmol/hr	19 785.17	23 429.00	24 331.94	452.16	354.57
	TOTAL DRY GAS	lb/hr	1 604	315 266	334 774	6 084	5 986
	H ₂ O	lb/hr	24 904	18	54		
	METHANOL	lb/hr	588 301	2 499	7 466	48	48
	TAR	lb/hr					
	OIL	lb/hr					
	NAPHTHA	lb/hr					
	CRUDE PHENOLS	lb/hr					
	ANHYDR. AMMONIA	lb/hr					
	SULPHUR	lb/hr					
	OTHERS	lb/hr					
	MINERALS	lb/hr					
	COAL	lb/hr					
	SULPHUR ORGANIC	lb/hr					
	NITROGEN ORGANIC	lb/hr					
	FATTY ACIDS	lb/hr					
	ACETONE	lb/hr					
	C ₆ +	lb/hr					
	TOTAL	lb/hr	614 809	317 783	342 294	6 133	5 917

MATERIAL BALANCE

BASE CASE I

STREAM N ^o		42	43	44	45	46
		Off gas to Unit 112	Steam to Unit 112	Unit 112 Pro- duct Gas	Fuel to Unit 114	Feed to Unit 114
CO ₂	44.010 lbmol/hr	263.99		567.08	0.76	85.00
H ₂ S	34.076 lbmol/hr					
COS	60.070 lbmol/hr					
C ₂ H ₄	28.054 lbmol/hr	1.39			0.05	3.12
CO	28.010 lbmol/hr	52.73		0.98	0.23	3.24
H ₂	2.016 lbmol/hr	139.88		271.42	0.66	275.95
CH ₄	16.043 lbmol/hr	713.62		15 459.37	5.18	15 566.75
C ₂ H ₆	30.070 lbmol/hr	14.13			0.40	34.83
N ₂ +INERT	28.013 lbmol/hr	10.84		308.97		308.67
O ₂	32.000 lbmol/hr					
TOTAL DRY GAS (1)	lbmol/hr	1 196.58		16 607.82	10.27	16 304.96
H ₂ O	18.016 lbmol/hr	2.02	2 400	13.32	0.17	17.43
METHANOL	32.042 lbmol/hr	156.50				
TOTAL WET GAS (1)	lbmol/hr	1 355.10		16 621.14	10.44	16 322.39
TOTAL DRY GAS	lb/hr	25 593		282 201		265 340
H ₂ O	lb/hr	36		240		314
METHANOL	lb/hr	5 014				
C ₃ H ₆	42.08 lbmol/hr				0.05	0.68
C ₃ H ₈	44.09 lbmol/hr				1.12	13.34
iC ₄ H ₁₀	58.12 lbmol/hr				0.84	7.99
1C ₄ H ₈	56.10 lbmol/hr				0.09	0.85
nC ₄ H ₁₀	58.12 lbmol/hr				0.20	1.77
1C ₅ H ₁₂	72.15 lbmol/hr				0.36	2.18
1C ₅ H ₁₀	70.13 lbmol/hr				0.06	0.30
nC ₅ H ₁₂	72.15 lbmol/hr				0.03	0.14
C ₆ +	(2) lbmol/hr				0.24	0.15
NAPHTHA	lb/hr					
TOTAL	lb/hr	30 644	43 238	282 441	311	265 654
(1) C ₃ + INCLUDED						
(2) MW OF C ₆ +						85.97

MATERIAL BALANCE

BASE CASE I

STREAM N°			47	48	49	50	51
			Stack gas from Unit 114	Unit 113 Off gas	SNG Product	Gasoline Product	Unit 112 Condensate
CO ₂	44.010	lbmol/hr	18.79	514.39	85.00		
H ₂ S	34.076	lbmol/hr					
COS	60.070	lbmol/hr					
C ₂ H ₄	28.054	lbmol/hr			3.12		
CO	28.010	lbmol/hr			3.24		
H ₂	2.016	lbmol/hr		0.29	275.95		
CH ₄	16.043	lbmol/hr		16.25	15 566.75		
C ₂ H ₆	30.070	lbmol/hr		0.04	34.83		
N ₂ +INERT	28.013	lbmol/hr	144.42	0.32	308.67		
O ₂	32.000	lbmol/hr	7.01				
TOTAL DRY GAS (1)		lbmol/hr	168.22	531.31	16 304.96		5 722
H ₂ O	18.016	lbmol/hr	31.55	41.19	1.33		
METHANOL	32.042	lbmol/hr					
TOTAL WET GAS (1)		lbmol/hr	199.77	572.50	16 306.29		
TOTAL DRY GAS		lb/hr	5 097	22 911	265 340		
H ₂ O		lb/hr	568	742	24		103 088
METHANOL		lb/hr					
C ₃ H ₆	42.08	lbmol/hr			0.68		
C ₃ H ₈	44.09	lbmol/hr			13.34	0.01	
iC ₄ H ₁₀	58.12	lbmol/hr		0.01	7.99	84.47	
1C ₄ H ₈	56.10	lbmol/hr		0.01	0.85	2.41	
nC ₄ H ₁₀	58.12	lbmol/hr			1.77	45.80	
iC ₅ H ₁₂	72.15	lbmol/hr			2.18	430.71	
1C ₅ H ₁₀	70.13	lbmol/hr			0.30	79.31	
nC ₅ H ₁₂	72.15	lbmol/hr			0.14	48.99	
C ₆ +	(2)	lbmol/hr			0.15	1 645.73	
NAPHTHA		lb/hr				15 541	
TOTAL		lb/hr	5 666	23 653	265 364	235 668	103 088
(1) C ₃ + INCLUDED							
(2) MW OF C ₆ +					85.97	104.65	

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MATERIAL BALANCE

BASE CASE I

STREAM N°			52	53	54	55	56
			Fuel to Unit 151	Stack gas fr. Unit 151	Light gas to SNG	Water make up to Unit 113	Steam from Unit 114
CO ₂	44.010	lbmol/hr	0.24	5.81	32.21		
H ₂ S	34.076	lbmol/hr					
COS	60.070	lbmol/hr					
C ₂ H ₄	28.054	lbmol/hr	0.01		3.12		
CO	28.010	lbmol/hr	0.07		2.26		
H ₂	2.016	lbmol/hr	0.21		4.82		
CH ₄	16.043	lbmol/hr	1.61		123.63		
C ₂ H ₆	30.070	lbmol/hr	0.12		34.87		
N ₂ + INERT	28.013	lbmol/hr		41.98	0.02		
O ₂	32.000	lbmol/hr		1.46			
TOTAL DRY GAS (1)		lbmol/hr	3.19	49.25	228.45		
H ₂ O	18.016	lbmol/hr	0.05	9.69	12.80	43.39	16.10
METHANOL	32.042	lbmol/hr					
TOTAL WET GAS (1)		lbmol/hr	3.24	58.94	230.36		
TOTAL DRY GAS		lb/hr	96	1 478	6 049		
H ₂ O		lb/hr	1	175	231	782	290
METHANOL		lb/hr					
C ₃ H ₆	42.08	lbmol/hr	0.02		0.68		
C ₃ H ₈	44.09	lbmol/hr	0.35		13.35		
iC ₄ H ₁₀	58.12	lbmol/hr	0.26		8.00		
1C ₄ H ₈	56.10	lbmol/hr	0.03		0.85		
nC ₄ H ₁₀	58.12	lbmol/hr	0.06		1.77		
iC ₅ H ₁₂	72.15	lbmol/hr	0.11		2.18		
1C ₅ H ₁₀	70.13	lbmol/hr	0.02		0.30		
nC ₅ H ₁₂	72.15	lbmol/hr	0.01		0.14		
C ₆ +	(2)	lbmol/hr	0.07		0.15		
NAPHTHA							
TOTAL			97	1 653	6 280	782	290
(1) C ₃ + INCLUDED							
(2) MW OF C ₆ +					85.97		

MATERIAL BALANCE

BASE CASE I

-245-

STREAM №			57	58	59	60	
			Light Gas from Unit 152	Condensate from Unit 113	Oil to Unit 122/123	Oil to Unit 124	
CO ₂	44.010 lbmol/hr		35.61	3.30			
H ₂ S	34.076 lbmol/hr						
COS	60.070 lbmol/hr						
C ₂ H ₄	28.054 lbmol/hr		3.52	0.40			
CO	28.010 lbmol/hr		2.34	0.08			
H ₂	2.016 lbmol/hr		4.90	0.08			
CH ₄	16.043 lbmol/hr		129.71	6.08			
C ₂ H ₆	30.070 lbmol/hr		39.97	5.10			
N ₂ + INERT	28.013 lbmol/hr		0.02				
O ₂	32.000 lbmol/hr						
TOTAL DRY GAS (1)	lbmol/hr		258.48	30.03			
H ₂ O	18.016 lbmol/hr		12.80				
METHANOL	32.042 lbmol/hr						
TOTAL WET GAS (1)	lbmol/hr		271.28				
TOTAL DRY GAS	lb/hr		7 336	1 287			
H ₂ O	lb/hr		231				
METHANOL	lb/hr						
C ₃ H ₆	42.08 lbmol/hr		0.88	0.20			
C ₃ H ₈	44.09 lbmol/hr		17.57	4.22			
iC ₄ H ₁₀	58.12 lbmol/hr		12.82	4.82			
1C ₄ H ₈	56.10 lbmol/hr		1.38	0.53			
nC ₄ H ₁₀	58.12 lbmol/hr		2.94	1.17			
iC ₅ H ₁₂	72.15 lbmol/hr		4.60	2.42			
1C ₅ H ₁₀	70.13 lbmol/hr		0.74	0.44			
nC ₅ H ₁₂	72.15 lbmol/hr		0.36	0.22			
C ₆ +	(2) lbmol/hr		1.12	0.97			
NAPHTHA	lb/hr						
OIL	lb/hr				35 903	15 808	
TOTAL	lb/hr		7 567	1 287	35 903	15 808	
(1) C ₃ + INCLUDED							
(2) MW OF C ₆ +			85.97	86.17			

MATERIAL BALANCE

BASE CASE I

-246-

STREAM N°		62	63	64	65	
		Unit 124 Stack gas	Units 122/123 Stack gas	SO ₂ from Unit 126	Stack gas to Atmosphere	
	CO ₂ 44.010 lbmol/hr	1 111.20	53 359.23		54 470.43	
	H ₂ S 34.076 lbmol/hr					
	SO ₂ 60.070 lbmol/hr	2.02	40.71	31.63	11.10	
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERTS 28.013 lbmol/hr	6 225.97	92 550.80		98 776.77	
	O ₂ 32.000 lbmol/hr	215.77	1 643.76		1 859.53	
	TOTAL dry gas lbmol/hr	7 554.96	147 594.50		155 117.83	
	H ₂ O 18.016 lbmol/hr	902.48	19 928.34		20 830.82	
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS lbmol/hr	8 457.44	167 522.84		175 948.65	
	TOTAL DRY GAS lb/hr	230 346	4 996 174	2 026	5 224 494	
	H ₂ O lb/hr	16 259	359 029		375 288	
	METHANOL lb/hr					
	TAR lb/hr					
	OIL lb/hr					
	NAPHTHA lb/hr					
	CRUDE PHENOLS lb/hr					
	ANHYDR. AMMONIA lb/hr					
	SULPHUR lb/hr					
	OTHERS lb/hr					
	MINERALS lb/hr					
	COAL lb/hr					
	SULPHUR ORGANIC lb/hr					
	NITROGEN ORGANIC lb/hr					
	FATTY ACIDS lb/hr					
	ACETONE lb/hr					
	C ₆ + lb/hr					
	TOTAL lb/hr	246 605	5 355 203	2 026	5 599 782	

MATERIAL BALANCE

BASE CASE I

STREAM №		66	67	68	69	70
		N ₂ to Atmosphere	Air to Plant	Air to Units 122/123	Stack gas to Unit 122	Air to Unit 121
	CO ₂ 44.010 lbmol/hr				592.42	
	H ₂ S 34.076 lbmol/hr				0.01	
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr				27.08	
	CH ₄ 16.043 lbmol/hr				22.28	
	C ₂ H ₆ 30.070 lbmol/hr				0.32	
	N ₂ + INERT 28.013 lbmol/hr	55 553.58	154 538.43	91 914.23	569.84	55 831.48
	O ₂ 32.000 lbmol/hr	781.28	41 079.84	24 432.90	21.75	14 841.28
	TOTAL DRY GAS (1) lbmol/hr	56 334.86	195 618.27	116 347.13	1 233.97	70 672.76
	H ₂ O 18.016 lbmol/hr	641.87	4 867.58	2 895.03	170.63	1 758.61
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS (1) lbmol/hr	56 976.73	200 485.58	119 242.16	1 404.60	72 431.37
	TOTAL DRY GAS lb/hr	581 223	5 643 638	3 356 645	43 170	2 038 923
	H ₂ O lb/hr	11 564	87 694	52 157	3 074	31 683
	METHANOL lb/hr					
	C ₃ H ₆ 42.08 lbmol/hr					
	C ₃ H ₈ 44.09 lbmol/hr				0.02	
	iC ₄ H ₁₀ 58.12 lbmol/hr				0.01	
	1C ₄ H ₈ 56.10 lbmol/hr				0.01	
	nC ₄ H ₁₀ 58.12 lbmol/hr				0.09	
	iC ₅ H ₁₂ 72.15 lbmol/hr					
	1C ₅ H ₁₀ 70.13 lbmol/hr					
	nC ₅ H ₁₂ 72.15 lbmol/hr				0.14	
	C ₆ + (2) lbmol/hr					
	NAPHTHA lb/hr					
	TOTAL lb/hr	592 787	5 731 332	3 408 802	46 244	2 070 606
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					

MATERIAL BALANCE

BASE CASE I

STREAM N°		71	72	73	74	75
		Water from Unit 121	Air to Unit 135	Air to Unit 114	Air to Unit 151	Unit 150 Acid Water
	CO ₂ 44.010 lbmol/hr					6.28
	H ₂ S 34.076 lbmol/hr					
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					0.11
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERT 28.013 lbmol/hr		9 437.32	144.42	41.98	
	O ₂ 32.000 lbmol/hr		2 508.66	38.39	11.16	
	TOTAL DRY GAS (1) lbmol/hr		11 945.98	182.81	53.14	6.39
	H ₂ O 18.016 lbmol/hr	1 116.73	297.25	4.55	1.32	19 627.00
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS (1) lbmol/hr		12 243.23	187.36	54.46	
	TOTAL DRY GAS lb/hr		344 645	5 274	1 533	279
	H ₂ O lb/hr	20 119	5 355	82	24	353 600
	METHANOL lb/hr					
	C ₃ H ₆ 42.08 lbmol/hr					
	C ₃ H ₈ 44.09 lbmol/hr					
	iC ₄ H ₁₀ 58.12 lbmol/hr					
	1C ₄ H ₈ 56.10 lbmol/hr					
	nC ₄ H ₁₀ 58.12 lbmol/hr					
	iC ₅ H ₁₂ 72.15 lbmol/hr					
	1C ₅ H ₁₀ 70.13 lbmol/hr					
	nC ₅ H ₁₂ 72.15 lbmol/hr					
	C ₆ + (2) lbmol/hr					
	NAPHTHA lb/hr					
	ACETONE lb/hr					1 377
	ACID lb/hr					1 379
	C ₆ + lb/hr					423
	TOTAL lb/hr	20 119	350 000	5 356	1 557	357 058
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					

MATERIAL BALANCE

BASE CASE I

STREAM N ^o		76	77	78		
		Water from Unit 151	Acid water to Unit 135	Steam to Unit 108		
	CO ₂ 44.010 lbmol/hr		6.28			
	H ₂ S 34.076 lbmol/hr	1.0	1.00			
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr		0.11			
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERTS 28.013 lbmol/hr					
	O ₂ 32.000 lbmol/hr					
	TOTAL DRY GAS lbmol/hr		7.39			
	H ₂ O 18.016 lbmol/hr	449.43	20 076.43	344.14		
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS lbmol/hr					
	TOTAL DRY GAS lb/hr	34	313			
	H ₂ O lb/hr	8 097	361 697	6 200		
	METHANOL lb/hr					
	TAR lb/hr					
	OIL lb/hr					
	NAPHTHA lb/hr					
	CRUDE PHENOLS lb/hr					
	ANHYDR. AMMONIA lb/hr	17	17			
	SULPHUR lb/hr					
	OTHERS lb/hr					
	MINERALS lb/hr					
	COAL lb/hr					
	SULPHUR ORGANIC lb/hr					
	NITROGEN ORGANIC lb/hr					
	FATTY ACIDS lb/hr		1 379			
	ACETONE lb/hr		1 377			
	C ₆ + lb/hr		423			
	TOTAL lb/hr	8 148	365 206	6 200		

MATERIAL BALANCE

BASE CASE I

STREAM N°			79	80		
			Unit 151 Purge Gas	Treated Water		
	CO ₂	44.010 lbmol/hr				
	H ₂ S	34.076 lbmol/hr	0.01			
	COS	60.070 lbmol/hr				
	C ₂ H ₄	28.054 lbmol/hr				
	CO	28.010 lbmol/hr				
	H ₂	2.016 lbmol/hr	26.79			
	CH ₄	16.043 lbmol/hr	6.03			
	C ₂ H ₆	30.070 lbmol/hr	0.28			
	N ₂ +INERT	28.013 lbmol/hr				
	O ₂	32.000 lbmol/hr				
	TOTAL DRY GAS (1)	lbmol/hr	33.36			
	H ₂ O	18.016 lbmol/hr	0.05	110 742.3		
	METHANOL	32.042 lbmol/hr				
	TOTAL WET GAS (1)	lbmol/hr	33.41			
	TOTAL DRY GAS	lb/hr	176			
	H ₂ O	lb/hr	1	1 994 893		
	METHANOL	lb/hr				
	C ₃ H ₆	42.08 lbmol/hr				
	C ₃ H ₈	44.09 lbmol/hr	0.02			
	iC ₄ H ₁₀	58.12 lbmol/hr				
	1C ₄ H ₈	56.10 lbmol/hr				
	nC ₄ H ₁₀	58.12 lbmol/hr	0.09			
	1C ₅ H ₁₂	72.15 lbmol/hr				
	1C ₅ H ₁₀	70.13 lbmol/hr				
	nC ₅ H ₁₂	72.15 lbmol/hr	0.14			
	C ₆ +	(2) lbmol/hr				
	NAPHTHA	lb/hr				
	TOTAL	lb/hr	177	1 994 893		
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					

MATERIAL BALANCE

BASE CASE I

-251-

STREAM N°		81	82	84	85.1	85.2
		BFW to Unit 106	Water from Unit 113	Air to Unit 124	Air to Unit 106	Air from Unit 106
CO ₂	44.010 lbmol/hr					
H ₂ S	34.076 lbmol/hr					
COS	60.070 lbmol/hr					
C ₂ H ₄	28.054 lbmol/hr				Air surplus + N ₂	Air surplus + N ₂
CO	28.010 lbmol/hr				not balanced	not balanced
H ₂	2.016 lbmol/hr					
CH ₄	16.043 lbmol/hr					
C ₂ H ₆	30.070 lbmol/hr					
N ₂ +INERTS	28.013 lbmol/hr			6 223.21		
O ₂	32.000 lbmol/hr			1 654.27	78.88	
TOTAL DRY GAS	lbmol/hr			7 877.48		
H ₂ O	18.016 lbmol/hr	2 130.88	13.30	196.00		
METHANOL	32.042 lbmol/hr					
TOTAL WET GAS	lbmol/hr			8 073.48		
TOTAL DRY GAS	lb/hr			227 267	2 524	
H ₂ O	lb/hr	38 390	240	3 531		
METHANOL	lb/hr					
TAR	lb/hr					
OIL	lb/hr					
NAPHTHA	lb/hr					
CRUDE PHENOLS	lb/hr					
ANHYDR. AMMONIA	lb/hr					
SULPHUR	lb/hr					
OTHERS	lb/hr					
MINERALS	lb/hr					
COAL	lb/hr					
SULPHUR ORGANIC	lb/hr					
NITROGEN ORGANIC	lb/hr					
FATTY ACIDS	lb/hr					
ACETONE	lb/hr					
C ₆ +	lb/hr					
TOTAL	lb/hr	38 390	240	230 798	2 524	0

MATERIAL BALANCE

BASE CASE I

STREAM N°		86	87	88	89	90
		BFW Make Up Water	BFW Blow Down	C.W. Make Up	Evaporation + Spray Losses from Unit 133	Unit 135 Air
	CO ₂ 44.010 lbmol/hr					8.01
	H ₂ S 34.076 lbmol/hr					1.88
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					0.11
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ + INERTS 28.013 lbmol/hr					9 437.32
	O ₂ 32.000 lbmol/hr					2 508.66
	TOTAL DRY GAS lbmol/hr					11 955.98
	H ₂ O 18.016 lbmol/hr	117 501.11	8 597.91	56 993.78	128 863.23	297.25
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS lbmol/hr					
	TOTAL DRY GAS lb/hr					345 064
	H ₂ O lb/hr	2 116 900	154 900	1 026 800	2 321 600	5 355
	METHANOL lb/hr					
	TAR lb/hr					
	OIL lb/hr					
	NAPHTHA lb/hr					
	CRUDE PHENOLS lb/hr					587
	ANHYDR. AMMONIA lb/hr					1 308
	SULPHUR lb/hr					
	OTHERS lb/hr					
	MINERALS lb/hr					
	COAL lb/hr					
	SULPHUR ORGANIC lb/hr					
	NITROGEN ORGANIC lb/hr					
	FATTY ACIDS lb/hr					4 178
	ACETONE lb/hr					1 377
	C ₆ + lb/hr					423
	TOTAL lb/hr	2 116 900	154 900	1 017 800	2 321 600	358 292

MATERIAL BALANCE

BASE CASE I

STREAM №		91	92	93	94	95
		CW Blow Down	Make Up Water to Unit 129	Sluice Vent	Steam to Unit 151	Deaerator Losses
	CO ₂ 44.010 lbmol/hr					
	H ₂ S 34.076 lbmol/hr			Air not balanced		
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERT 28.013 lbmol/hr					
	O ₂ 32.000 lbmol/hr					
	TOTAL DRY GAS (1) lbmol/hr					
	H ₂ O 18.016 lbmol/hr	36 334.37	2 525.53	1 415.41	394.09	555.06
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS (1) lbmol/hr					
	TOTAL DRY GAS lb/hr					
	H ₂ O lb/hr	654 600	45 500	25 500	7 100	10 000
	METHANOL lb/hr					
	C ₃ H ₆ 42.08 lbmol/hr					
	C ₃ H ₈ 44.09 lbmol/hr					
	iC ₄ H ₁₀ 58.12 lbmol/hr					
	1C ₄ H ₈ 56.10 lbmol/hr					
	nC ₄ H ₁₀ 58.12 lbmol/hr					
	iC ₅ H ₁₂ 72.15 lbmol/hr					
	1C ₅ H ₁₀ 70.13 lbmol/hr					
	nC ₅ H ₁₂ 72.15 lbmol/hr					
	C ₆ + (2) lbmol/hr					
	NAPHTHA lb/hr					
	TOTAL lb/hr	654 600	45 500	25 500	7 100	10 000
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					

MATERIAL BALANCE

BASE CASE I

STREAM N°			96	97	98	99	100
			Unit 150 Rege- generation Purge Gas	Unit 150 Off Gas	Unit 150HC Product	Stabilized ga- soline from Unit 152	Naphtha from Unit 151
	CO ₂	44.010 lbmol/hr		3.16	32.31		
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr		0.19	3.12		
	CO	28.010 lbmol/hr		0.95	2.26		
	H ₂	2.016 lbmol/hr		2.77	4.82		
	CH ₄	16.043 lbmol/hr		21.51	123.63		
	C ₂ H ₆	30.070 lbmol/hr		1.63	35.85		
	N ₂ +INERT	28.013 lbmol/hr		0.01	0.02		
	O ₂	32.000 lbmol/hr					
	TOTAL DRY GAS (1)	lbmol/hr		42.67	3 194.84		
	H ₂ O	18.016 lbmol/hr		0.71	12.80		
	METHANOL	32.042 lbmol/hr					
	TOTAL WET GAS (1)	lbmol/hr		43.38	3 207.64		
	TOTAL DRY GAS	lb/hr		1 282			
	H ₂ O	lb/hr		13			
	METHANOL	lb/hr					
	C ₃ H ₆	42.08 lbmol/hr		0.23	11.54		
	C ₃ H ₈	44.09 lbmol/hr		4.65	266.32	0.01	
	iC ₄ H ₁₀	58.12 lbmol/hr		3.49	392.66	8.43	
	1C ₄ H ₈	56.10 lbmol/hr		0.39	49.25	2.41	
	nC ₄ H ₁₀	58.12 lbmol/hr		0.83	121.42	13.12	1.98
	iC ₅ H ₁₂	72.15 lbmol/hr		1.47	431.08	421.50	158.43
	1C ₅ H ₁₀	70.13 lbmol/hr		0.26	79.71	78.80	
	nC ₅ H ₁₂	72.15 lbmol/hr		0.14	49.13	48.96	
	C ₆ + (2)	lbmol/hr		0.99	1 591.72	1 591.57	
	COKE MWT 12.82	lb/hr	300				
	NAPHTHA	lb/hr					15 541
	TOTAL	lb/hr	300	1 295	256 156	207 023	15 541
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +			90.14	104.40	104.40	

MATERIAL BALANCE

BASE CASE I

STREAM №		101	102	103	104	105
		Fuel to Unit 150	Propane LPG Product	Mixed Butane Product	Alkylate to Unit 154	Feed to Unit 153
	CO ₂ 44.010 lbmol/hr	2.16				
	H ₂ S 34.076 lbmol/hr					
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr	0.13				
	CO 28.010 lbmol/hr	0.65				
	H ₂ 2.016 lbmol/hr	1.90				
	CH ₄ 16.043 lbmol/hr	14.72				
	C ₂ H ₆ 30.070 lbmol/hr	1.11	0.98			0.98
	N ₂ +INERT 28.013 lbmol/hr	0.01				
	O ₂ 32.000 lbmol/hr					
	TOTAL DRY GAS (1) lbmol/hr	29.21	259.81	314.01		801.59
	H ₂ O 18.016 lbmol/hr	0.49				
	METHANOL 32.042 lbmol/hr	29.70				
	TOTAL WET GAS (1) lbmol/hr					
	TOTAL DRY GAS lb/hr	878	11 501	18 277		
	H ₂ O lb/hr	9				
	METHANOL lb/hr					
	C ₃ H ₆ 42.08 lbmol/hr	0.16				10.86
	C ₃ H ₈ 44.09 lbmol/hr	3.18	254.58			252.96
	iC ₄ H ₁₀ 58.12 lbmol/hr	2.39	4.24	238.11	2.35	376.23
	1C ₄ H ₈ 56.10 lbmol/hr	0.27	0.01			45.99
	nC ₄ H ₁₀ 58.12 lbmol/hr	0.57		73.95	9.80	106.53
	iC ₅ H ₁₂ 72.15 lbmol/hr	1.00		1.85	8.64	7.40
	1C ₅ H ₁₀ 70.13 lbmol/hr	0.18		0.10	0.48	0.61
	nC ₅ H ₁₂ 72.15 lbmol/hr	0.10			0.03	0.03
	C ₆ + (2) lbmol/hr	0.68			54.16	
	NAPHTHA lb/hr					
	TOTAL lb/hr	887	11 501	18 277	7 448.	42 856
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +	90.14			112.28	

MATERIAL BALANCE

BASE CASE I

STREAM No		107	108	109	110	111
		Wash Water to Unit 151	Stack Gas from Unit 150	Air to Unit 150	Mixed Butanes to Unit 154	Unit 132 Blow Down
	CO ₂ 44.010 lbmol/hr		53.43			
	H ₂ S 34.076 lbmol/hr					
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERT 28.013 lbmol/hr		383.12	383.11		
	O ₂ 32.000 lbmol/hr		13.28	101.84		
	TOTAL DRY GAS (1) lbmol/hr		449.83	484.95	97.17	
	H ₂ O 18.016 lbmol/hr	48.29	88.15	12.07		959.87
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS (1) lbmol/hr		537.98	497.02		
	TOTAL DRY GAS lb/hr		13 508	13 991	5 656	
	H ₂ O lb/hr	870	1 588	217		17 293
	METHANOL lb/hr					
	C ₃ H ₆ 42.08 lbmol/hr					
	C ₃ H ₈ 44.09 lbmol/hr					
	iC ₄ H ₁₀ 58.12 lbmol/hr				73.69	
	1C ₄ H ₈ 56.10 lbmol/hr					
	nC ₄ H ₁₀ 58.12 lbmol/hr				22.88	
	iC ₅ H ₁₂ 72.15 lbmol/hr				0.57	
	1C ₅ H ₁₀ 70.13 lbmol/hr				0.03	
	nC ₅ H ₁₂ 72.15 lbmol/hr					
	C ₆ + (2) lbmol/hr					
	NAPHTHA lb/hr					
	TOTAL lb/hr	870	15 096	14 208	5 656	17 293
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					

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APPENDIX B

MATERIAL BALANCE - BASE CASE II

MATERIAL BALANCE

BASE CASE II

STREAM N ^o		1	2	3	4	5
		Mined Coal	Fine Coal to Unit 222	Well sized coal	Oxygen	MP Steam
CO ₂	44.010 lbmol/hr					
H ₂ S	34.076 lbmol/hr					
COS	60.070 lbmol/hr					
C ₂ H ₄	28.054 lbmol/hr					
CO	28.010 lbmol/hr					
H ₂	2.016 lbmol/hr					
CH ₄	16.043 lbmol/hr					
C ₂ H ₆	30.070 lbmol/hr					
N ₂ + INERTS	28.013 lbmol/hr				277.9	
O ₂	32.000 lbmol/hr				14 060	
TOTAL DRY GAS	lbmol/hr				14 337.9	
H ₂ O	18.016 lbmol/hr					94 379.2
METHANOL	32.042 lbmol/hr					
TOTAL WET GAS	lbmol/hr					94 379.2
TOTAL DRY GAS	lb/hr				457 697	
H ₂ O	lb/hr	648 487	116 097	532 390		1700 335
METHANOL	lb/hr					
TAR	lb/hr					
OIL	lb/hr					
NAPHTHA	lb/hr					
CRUDE PHENOLS	lb/hr					
ANHYDR. AMMONIA	lb/hr					
SULPHUR	lb/hr					
OTHERS	lb/hr					
MINERALS	lb/hr	118 117	21 146	96 971		
COAL	lb/hr	1549 420	277 457	1271 963		
SULPHUR ORGANIC	lb/hr					
NITROGEN ORGANIC	lb/hr					
FATTY ACIDS	lb/hr					
ACETONE	lb/hr					
C ₆ +	lb/hr					
TOTAL	lb/hr	2316 024	414 700	1901 324	457 697	1700 335

MATERIAL BALANCE

BASE CASE II

STREAM N°		5.1	6	6.1	6.2	6.3
		BFW to Unit 201	Ash to Disposal	Ash from Unit 201	Ash from Unit 222	Make Up Water to Unit 229
	CO ₂ 44.010 lbmol/hr					
	H ₂ S 34.076 lbmol/hr					
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERTS 28.013 lbmol/hr					
	O ₂ 32.000 lbmol/hr					
	TOTAL DRY GAS 1 lbmol/hr					
	H ₂ O 18.016 lbmol/hr	10 591	1 110.12	1 110.12		2 525.53
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS 1 lbmol/hr	10 591				
	TOTAL DRY GAS 1b/hr					
	H ₂ O 1b/hr	190 811	20 000	20 000		45 500
	METHANOL 1b/hr					
	TAR 1b/hr					
	OIL 1b/hr					
	NAPHTHA 1b/hr					
	CRUDE PHENOLS 1b/hr					
	ANHYDR. AMMONIA 1b/hr					
	SULPHUR 1b/hr		66	66		
	OTHERS 1b/hr					
	MINERALS 1b/hr		118 121	96 971	21 150	
	COAL 1b/hr		7 038			
	SULPHUR ORGANIC 1b/hr					
	NITROGEN ORGANIC 1b/hr					
	FATTY ACIDS 1b/hr					
	ACETONE 1b/hr					
	C ₆ + 1b/hr					
	SO ₂ 1)		2 577			
	TOTAL 1b/hr	190 811	147 802	124 075	21 150	45 500
	1) mineralized, but balanced as SO ₂					

MATERIAL BALANCE

BASE CASE II

STREAM N°			7	8	9	10	11
			Dusty Gas liquor	Unit 203 Gas liquor	Unit 204 Gas liquor	Unit 205 Gas liquor	Unit 207 Feed
CO ₂	44.010	lbmol/hr	1.73				1.73
H ₂ S	34.076	lbmol/hr	0.88				0.88
COS	60.070	lbmol/hr					
C ₂ H ₄	28.054	lbmol/hr					
CO	28.010	lbmol/hr					
H ₂	2.016	lbmol/hr					
CH ₄	16.043	lbmol/hr					
C ₂ H ₆	30.070	lbmol/hr					
N ₂ +INERTS	28.013	lbmol/hr					
O ₂	32.000	lbmol/hr					
TOTAL DRY GAS		lbmol/hr	2.61				2.61
H ₂ O	18.016	lbmol/hr	54541.8	18895.0	12041.4	134.3	85612.57
METHANOL	32.042	lbmol/hr					
TOTAL WET GAS		lbmol/hr					
TOTAL DRY GAS		lb/hr	106				106
H ₂ O		lb/hr	982625	340413	216938	2420	1542396
METHANOL		lb/hr					
TAR		lb/hr	31951	3604	836		36391
OIL		lb/hr	13444	31221	7246		51911
NAPHTHA		lb/hr					
CRUDE PHENOLS		lb/hr	9959	4248	986		15193
ANHYDR. AMMONIA		lb/hr	1734	6556	1522		9812
SULPHUR		lb/hr					
OTHERS		lb/hr					
MINERALS		lb/hr					
COAL		lb/hr	4113				4113
SULPHUR ORGANIC		lb/hr					
NITROGEN ORGANIC		lb/hr					
FATTY ACIDS		lb/hr	1770	835	194		2799
ACETONE		lb/hr					
C ₆ +		lb/hr					
TOTAL		lb/hr	1045702	386877	227722	2420	1662721

MATERIAL BALANCE

BASE CASE II

STREAM N°		12	13	14	15	16
		Tar to Unit 222/223	Oil from Unit 207	Unit 208 Feed	Phenols to Boiler 222/223	Unit 209 Feed
	CO ₂ 44.010 lbmol/hr			1.73		1.73
	H ₂ S 34.076 lbmol/hr			0.88		0.88
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERTS 28.013 lbmol/hr					
	O ₂ 32.000 lbmol/hr					
	TOTAL DRY GAS lbmol/hr			2.61		2.61
	H ₂ O 18.016 lbmol/hr			85 482.29		85 801.34
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS lbmol/hr					
	TOTAL DRY GAS lb/hr			106		106
	H ₂ O lb/hr	2 347		1540049	452	1 545 797
	METHANOL lb/hr					
	TAR lb/hr	36 391				
	OIL lb/hr		51 711	200	200	
	NAPHTHA lb/hr					
	CRUDE PHENOLS lb/hr			15 193	14 606	587
	ANHYDR. AMMONIA lb/hr			9 812		9 812
	SULPHUR lb/hr					
	OTHERS lb/hr					
	MINERALS lb/hr					
	COAL lb/hr	4 113				
	SULPHUR ORGANIC lb/hr					
	NITROGEN ORGANIC lb/hr					
	FATTY ACIDS lb/hr			2 799		2 799
	ACETONE lb/hr					
	C ₆ + lb/hr					
	TOTAL lb/hr	42 851	51 711	1568 159	15 258	1 559 101

MATERIAL BALANCE

BASE CASE II

STREAM N°		17	18	18.1	18.2	18.3
		NH ₃ Product	Water to Unit 235	Water from Unit 209	Water from Unit 205	Unit 205 Make up water
	CO ₂ 44.010 lbmol/hr		1.73	1.73		
	H ₂ S 34.076 lbmol/hr		0.88	0.88		
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERTS 28.013 lbmol/hr					
	O ₂ 32.000 lbmol/hr					
	TOTAL DRY GAS lbmol/hr		2.61	2.61		
	H ₂ O 18.016 lbmol/hr		89 535.8	85 799.12	3 736.68	3 736.68
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS lbmol/hr					
	TOTAL DRY GAS lb/hr		106	106		
	H ₂ O lb/hr	40	1613077	1 545757	67 320	67 320
	METHANOL lb/hr					
	TAR lb/hr					
	OIL lb/hr					
	NAPHTHA lb/hr					
	CRUDE PHENOLS lb/hr		587	587		
	ANHYDR. AMMONIA lb/hr	8 521	1 291	1 291		
	SULPHUR lb/hr					
	OTHERS lb/hr					
	MINERALS lb/hr					
	COAL lb/hr					
	SULPHUR ORGANIC lb/hr					
	NITROGEN ORGANIC lb/hr					
	FATTY ACIDS lb/hr		2 799	2 799		
	ACETONE lb/hr					
	C ₆ + lb/hr					
	TOTAL lb/hr	8 561	1 617 860	1 550 540	67 320	67 320

MATERIAL BALANCE

BASE CASE II

STREAM №			19	20	21	22	23
			Raw gas from Unit 201	Raw gas to Unit 203	Raw gas from Unit 203	Raw gas to Unit 202	Shift gas to Unit 204
CO ₂	44.010 lbmol/hr		33 722.0	27 369.5	27 369.5	6 352.5	9 176.8
H ₂ S	34.076 lbmol/hr		157.3	127.7	127.7	29.6	29.6
COS	60.070 lbmol/hr		2	1.6	1.6	0.4	0.4
C ₂ H ₄	28.054 lbmol/hr		85.5	69.4	69.4	16.1	16.1
CO	28.010 lbmol/hr		21 458.5	17 416.1	17 416.1	4 042.4	1 218.1
H ₂	2.016 lbmol/hr		44 332	35 972.6	35 972.6	8 359.4	11 183.7
CH ₄	16.043 lbmol/hr		12 866	10 442.3	10 442.3	2 423.7	2 423.7
C ₂ H ₆	30.070 lbmol/hr		614	498.3	498.3	115.7	115.7
N ₂ +INERTS	28.013 lbmol/hr		310.3	251.8	251.8	58.5	58.5
O ₂	32.000 lbmol/hr						
TOTAL DRY GAS	lbmol/hr		113 547.6	92 149.3	92 149.3	21 398.3	24 222.6
H ₂ O	18.016 lbmol/hr		79 122.8	64 217.6	137.3	14 905.2	12 080.8
METHANOL	32.042 lbmol/hr						
TOTAL WET GAS	lbmol/hr		192 670.4	156 366.9	92 286.6	36 303.5	36 303.4
TOTAL DRY GAS	lb/hr		2 415 949	1 960 831	1 960 831	455 118	506 001
H ₂ O	lb/hr		1 425 476	1 156 945	2 474	268 531	217 648
METHANOL	lb/hr						
TAR	lb/hr		4 440	3 604		836	836
OIL	lb/hr		38 467	31 221		7 246	7 246
NAPHTHA	lb/hr		15 675	12 722	12 722	2 953	2 953
CRUDE PHENOLS	lb/hr		5 234	4 248		986	986
ANHYDR. AMMONIA	lb/hr		8 078	6 556		1 522	1 522
SULPHUR	lb/hr						
OTHERS	lb/hr						
MINERALS	lb/hr						
COAL	lb/hr						
SULPHUR ORGANIC	lb/hr		75	61	61	14	14
NITROGEN ORGANIC	lb/hr		14	11	11	3	3
FATTY ACIDS	lb/hr		1029	835		194	194
ACETONE	lb/hr						
C ₆ +	lb/hr						
TOTAL	lb/hr		3 914 437	3 177 034	1 976 099	737 403	737 403

MATERIAL BALANCE

BASE CASE II

STREAM №			24	25	26	27	
			Shift gas from Unit 204	Feed to Unit 205	Naphtha to Unit 251	Acid gas to Unit 206	
	CO ₂	44.010 lbmol/hr	9 176.8	36 546.3		33 646.3	
	H ₂ S	34.076 lbmol/hr	29.6	157.3		157.3	
	COS	60.070 lbmol/hr	0.4	2		2	
	C ₂ H ₄	28.054 lbmol/hr	16.1	85.5		64.5	
	CO	28.010 lbmol/hr	1 218.1	18 634.2		134.2	
	H ₂	2.016 lbmol/hr	11 183.7	47 156.3		171.3	
	CH ₄	16.043 lbmol/hr	2 423.7	12 866		406	
	C ₂ H ₆	30.070 lbmol/hr	115.7	614		337	
	N ₂ +INERTS	28.013 lbmol/hr	58.5	310.3		1.3	
	O ₂	32.000 lbmol/hr				14.45	
	TOTAL DRY GAS	lbmol/hr	24 222.6	116 371.9		34 919.9	
	H ₂ O	18.016 lbmol/hr	39.4	176.7		42.4	
	METHANOL	32.042 lbmol/hr					
	TOTAL WET GAS	lbmol/hr	24 262.0	116 548.6		34 976.75	
	TOTAL DRY GAS	lb/hr	506 001	2 466 832		1 508 846	
	H ₂ O	lb/hr	710	3 184		764	
	METHANOL	lb/hr			158	463	
	TAR	lb/hr					
	OIL	lb/hr					
	NAPHTHA	lb/hr	2 953	15 675	15 675		
	CRUDE PHENOLS	lb/hr					
	ANHYDR. AMMONIA	lb/hr					
	SULPHUR	lb/hr					
	OTHERS	lb/hr					
	MINERALS	lb/hr					
	COAL	lb/hr					
	SULPHUR ORGANIC	lb/hr	14	75	75		
	NITROGEN ORGANIC	lb/hr	3	14	14		
	FATTY ACIDS	lb/hr					
	ACETONE	lb/hr					
	C ₆ +	lb/hr					
	TOTAL	lb/hr	509 681	2 485 780	15 922	1 510 073	

MATERIAL BALANCE

BASE CASE II

-267-

STREAM N°			28	29	30	31	
			Unit 206 Off Gas	Sulphur Prod.	Feed to Unit 250	Unit 201 Quench water	
	CO ₂	44.010 lbmol/hr	33 646.3		2 900		
	H ₂ S	34.076 lbmol/hr	2.02				
	COS	60.070 lbmol/hr	2				
	C ₂ H ₄	28.054 lbmol/hr	64.5		21		
	CO	28.010 lbmol/hr	134.2		18 500		
	H ₂	2.016 lbmol/hr	174.13		46 985		
	CH ₄	16.043 lbmol/hr	410.64		12 460		
	C ₂ H ₆	30.070 lbmol/hr	337.82		277		
	N ₂ +INERT	28.013 lbmol/hr	1.30		309		
	O ₂	32.000 lbmol/hr					
	TOTAL DRY GAS (1)	lbmol/hr	34 773.99		81 452		
	H ₂ O	18.016 lbmol/hr	2 330.82			45 185.3	
	METHANOL	32.042 lbmol/hr	14.45		1.31		
	TOTAL WET GAS (1)	lbmol/hr	37 119.26		81 453.31		
	TOTAL DRY GAS	lb/hr	1 503 723		957 986		
	H ₂ O	lb/hr	41 992			814 058	
	METHANOL	lb/hr	463		42		
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr	0.18				
	iC ₄ H ₁₀	58.12 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	nC ₄ H ₁₀	58.12 lbmol/hr	0.83				
	iC ₅ H ₁₂	72.15 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	nC ₅ H ₁₂	72.15 lbmol/hr	0.07				
	C ₆ +	(2) lbmol/hr					
	NAPHTHA	lb/hr					
	SULPHUR			5 051			
	TOTAL	lb/hr	1 546 178	5 051	958 028	814 058	
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						

MATERIAL BALANCE

BASE CASE II

STREAM N°			32	33	34		
			Inc. gas to unit 222	Unit 211 Productgas	H ₂ S from Unit 251		
CO ₂	44.010	lbmol/hr	36 006.31	3.45	0.29		
H ₂ S	34.076	lbmol/hr	2.03		2.25		
COS	60.070	lbmol/hr	2.00				
C ₂ H ₄	28.054	lbmol/hr	65.00				
CO	28.010	lbmol/hr	134.24				
H ₂	2.016	lbmol/hr	203.62	1 131.99	2.83		
CH ₄	16.043	lbmol/hr	464.74	13.79	4.55		
C ₂ H ₆	30.070	lbmol/hr	339.30		0.82		
N ₂ +INERTS	28.013	lbmol/hr	5236.06				
O ₂	32.000	lbmol/hr	176.67				
TOTAL DRY FLOW (1)		lbmol/hr	42631.76	1 149.23	11.82		
H ₂ O	18.016	lbmol/hr	3683.94				
METHANOL	32.042	lbmol/hr	14.45				
TOTAL WET FLOW (1)		lbmol/hr	46330.15				
C ₃ H ₈	42.08	lbmol/hr	0.50				
C ₃ H ₆	44.09	lbmol/hr	0.25		0.18		
iC ₄ H ₈	56.10	lbmol/hr					
iC ₄ H ₆	56.10	lbmol/hr					
iC ₄ H ₁₀	58.12	lbmol/hr					
nC ₄ H ₁₀	58.12	lbmol/hr	0.83		0.83		
iC ₅ H ₁₀	70.13	lbmol/hr					
iC ₅ H ₁₂	70.13	lbmol/hr					
iC ₅ H ₁₄	72.15	lbmol/hr					
nC ₅ H ₁₂	72.15	lbmol/hr	0.21		0.07		
iC ₆ H ₁₂	84.16	lbmol/hr					
iC ₆ H ₁₄	84.16	lbmol/hr					
iC ₆ H ₁₆	86.17	lbmol/hr					
nC ₆ H ₁₄	86.17	lbmol/hr					
C ₆ +	(2)	lbmol/hr					
C ₇ +	(3)	lbmol/hr					
ALCOHOLS	46.61	lbmol/hr					
TOTAL		lb/hr	1 827 738	2 655	254		
(1) C ₃ + INCLUDED							
(2) MW OF C ₆ +							
(3) MW OF C ₇ +							

MATERIAL BALANCE

BASE CASE II

[illegible]

MATERIAL BALANCE

BASE CASE II

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STREAM N°			37	38	39	40	41
			UNIT 210 OFF STREAMS	UNIT 210 PRODUCT GAS	UNIT 212 FEED	UNIT 211 FEED	UNIT 211 OFF GAS
	CO ₂	44.010 lbmol/hr	1 070.00	2 014.66	2 011.21	409.35	405.90
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr	194.13	184.00	184.00	37.39	37.39
	CO	28.010 lbmol/hr	9.00	474.04	474.04	96.32	96.32
	H ₂	2.016 lbmol/hr	49.00	9 285.26	8 153.27	1 886.65	754.66
	CH ₄	16.043 lbmol/hr	1 010.00	13 217.78	13 203.99	2 685.69	2 671.90
	C ₂ H ₆	30.070 lbmol/hr	460.76	350.00	350.00	71.12	71.12
	N ₂ +INERTS	28.013 lbmol/hr	5.00	303.79	303.79	61.73	61.73
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	4 507.48	25 829.53	24 680.30	5 248.25	4 099.02
	H ₂ O	18.016 lbmol/hr	117.26				
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr	4 624.74				
	C ₃ H ₆	42.08 lbmol/hr	728.15				
	C ₃ H ₈	44.09 lbmol/hr	121.20				
	iC ₄ H ₈	56.10 lbmol/hr	352.49				
	iC ₄ H ₈	56.10 lbmol/hr	18.78				
	iC ₄ H ₁₀	58.12 lbmol/hr	2.43				
	nC ₄ H ₁₀	58.12 lbmol/hr	45.29				
	iC ₅ H ₁₀	70.13 lbmol/hr	139.52				
	iC ₅ H ₁₀	70.13 lbmol/hr	47.68				
	iC ₅ H ₁₂	72.15 lbmol/hr	8.06				
	nC ₅ H ₁₂	72.15 lbmol/hr	23.09				
	iC ₆ H ₁₂	84.16 lbmol/hr	63.09				
	iC ₆ H ₁₂	84.16 lbmol/hr	28.33				
	iC ₆ H ₁₄	86.17 lbmol/hr	6.66				
	nC ₆ H ₁₄	86.17 lbmol/hr	13.92				
	C ₆ +	(2) lbmol/hr					
	C ₇ +	(3) lbmol/hr	60.90	5.00	5.00	0.20	0.20
	ALCOHOLS	46.61 lbmol/hr	50.00				
	TOTAL	lb/hr	178 420	357 411	354 756	72 540	69 885
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +			100.00	100.00	100.00	100.00

MATERIAL BALANCE

BASE CASE II

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STREAM N°			42	43	44	45	46
			H ₂ TO UNIT 25	STEAM TO UNIT 212	UNIT 212 PRODUCT GAS	FUEL TO UNIT 214	UNIT 114 FEED
CO ₂	44.010	lbmol/hr	0.10		680.17	0.89	85.00
H ₂ S	34.076	lbmol/hr					
COS	60.070	lbmol/hr					
C ₂ H ₄	28.054	lbmol/hr				0.16	187.58
CO	28.010	lbmol/hr			4.26	0.01	13.14
H ₂	2.016	lbmol/hr	96.22		671.77	1.22	718.83
CH ₄	16.043	lbmol/hr	1.37		16 107.81	1.89	17 055.10
C ₂ H ₆	30.070	lbmol/hr				2.37	445.37
N ₂ +INERTS	28.013	lbmol/hr			303.79		307.97
O ₂	32.000	lbmol/hr					
TOTAL DRY FLOW (1)		lbmol/hr	97.69		17 767.80	8.60	19 021.79
H ₂ O	18.016	lbmol/hr		3 099.86	81.43	0.20	126.90
METHANOL	32.042	lbmol/hr					
TOTAL WET FLOW (1)		lbmol/hr			17 849.23	8.80	19 148.69
C ₃ H ₆	42.08	lbmol/hr				0.16	189.49
C ₃ H ₈	44.09	lbmol/hr				0.87	18.25
iC ₄ H ₈	56.10	lbmol/hr					0.24
iC ₄ H ₈	56.10	lbmol/hr					
iC ₄ H ₁₀	58.12	lbmol/hr				0.19	0.06
nC ₄ H ₁₀	58.12	lbmol/hr				0.13	0.03
iC ₅ H ₁₀	70.13	lbmol/hr					0.04
iC ₅ H ₁₀	70.13	lbmol/hr					
iC ₅ H ₁₂	72.15	lbmol/hr				0.04	
nC ₅ H ₁₂	72.15	lbmol/hr				0.06	0.01
iC ₆ H ₁₂	84.16	lbmol/hr					0.01
iC ₆ H ₁₂	84.16	lbmol/hr					
iC ₆ H ₁₄	86.17	lbmol/hr					0.01
nC ₆ H ₁₄	86.17	lbmol/hr				0.01	0.01
C ₆ +	(2)	lbmol/hr					
C ₇ +	(3)	lbmol/hr					0.65
ALCOHOLS	46.61	lbmol/hr					
TOTAL		lb/hr	220	55 847	299 803	225	317 644
(1) C ₃ + INCLUDED							
(2) MW OF C ₆ +							
(3) MW OF C ₇ +							153.24

MATERIAL BALANCE

BASE CASE II

STREAM N°			47	48	49	50	51
			STACK GAS FR. UNIT 214	UNIT 113 OFFGAS	SNG PRODUCT	GASOLINE PRODUCT	UNIT 112 CONDENSATE
CO ₂	44.010	lbmol/hr	12.78	1 635.42	85.00		
H ₂ S	34.076	lbmol/hr					
COS	60.070	lbmol/hr					
C ₂ H ₄	28.054	lbmol/hr		0.50	187.58		
CO	28.010	lbmol/hr		0.04	13.14		
H ₂	2.016	lbmol/hr		2.70	718.83		
CH ₄	16.043	lbmol/hr		48.18	17 055.10		
C ₂ H ₆	30.070	lbmol/hr		1.20	445.37		
N ₂ + INERTS	28.013	lbmol/hr	93.11	0.80	307.97		
O ₂	32.000	lbmol/hr	3.23				
TOTAL DRY FLOW (1)		lbmol/hr	109.12	1 689.39	19 021.79		
H ₂ O	18.016	lbmol/hr	22.39	101.57	1.50		6 150.09
METHANOL	32.042	lbmol/hr					
TOTAL WET FLOW (1)		lbmol/hr	131.51	1 790.96	19 023.29		
C ₃ H ₆	42.08	lbmol/hr		0.50	189.49		
C ₃ H ₈	44.09	lbmol/hr		0.05	18.25		
iC ₄ H ₈	56.10	lbmol/hr			0.24	4.68	
iC ₄ H ₁₀	56.10	lbmol/hr					
iC ₄ H ₁₂	58.12	lbmol/hr			0.06	1.32	
nC ₄ H ₁₀	58.12	lbmol/hr			0.03	89.05	
iC ₅ H ₁₀	70.13	lbmol/hr			0.04	7.55	
iC ₅ H ₁₂	70.13	lbmol/hr					
iC ₅ H ₁₄	72.15	lbmol/hr				305.63	
nC ₅ H ₁₂	72.15	lbmol/hr			0.01	55.53	
iC ₆ H ₁₂	84.16	lbmol/hr			0.01		
iC ₆ H ₁₄	84.16	lbmol/hr				173.67	
iC ₆ H ₁₆	86.17	lbmol/hr			0.01	30.72	
nC ₆ H ₁₄	86.17	lbmol/hr			0.01		
C ₆ +	(2)	lbmol/hr					
C ₇ +	(3)	lbmol/hr			0.65		
NAPHTHA	97.36	lbmol/hr				158.43	
ALKYLATE	102.51	lbmol/hr				332.07	
CAT POLY GASOLINE		lbmol/hr				350.88	
TOTAL	118.145		3 677	74 680	315 385	140 918	110 800
(1) C ₃ + INCLUDED							
(2) MW OF C ₆ +							
(3) MW OF C ₇ +					153.24		

MATERIAL BALANCE

BASE CASE II

STREAM N ^o			55	56	57		
			WATER MAKE UP TO UNIT 213	STEAM FROM UNIT 214	LIGHT GAS TO UNIT 213		
	CO ₂	44.010 lbmol/hr			1.040.25		
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr			188.08		
	CO	28.010 lbmol/hr			8.92		
	H ₂	2.016 lbmol/hr			49.76		
	CH ₄	16.043 lbmol/hr			995.47		
	C ₂ H ₆	30.070 lbmol/hr			446.57		
	N ₂ +INERTS	28.013 lbmol/hr			4.98		
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr			2 943.32		
	H ₂ O	18.016 lbmol/hr	183.56	125.4	111.79		
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr			3 055.11		
	C ₃ H ₆	42.08 lbmol/hr			189.99		
	C ₃ H ₈	44.09 lbmol/hr			18.30		
	1C ₄ H ₈	56.10 lbmol/hr			0.24		
	1C ₄ H ₈	56.10 lbmol/hr					
	1C ₄ H ₁₀	58.12 lbmol/hr			0.06		
	nC ₄ H ₁₀	58.12 lbmol/hr			0.03		
	1C ₅ H ₁₀	70.13 lbmol/hr			0.04		
	1C ₅ H ₁₀	70.13 lbmol/hr					
	1C ₅ H ₁₂	72.15 lbmol/hr					
	nC ₅ H ₁₂	72.15 lbmol/hr			0.01		
	1C ₆ H ₁₂	84.16 lbmol/hr			0.01		
	1C ₆ H ₁₂	84.16 lbmol/hr					
	1C ₆ H ₁₄	86.17 lbmol/hr			0.01		
	nC ₆ H ₁₄	86.17 lbmol/hr			0.01		
	C ₆ +	(2) lbmol/hr					
	C ₇ +	(3) lbmol/hr			0.65		
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	3 307	2 259	91 879		
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +				153.24		

MATERIAL BALANCE

BASE CASE II

STREAM N°		59	60			
		OIL TO UNIT 222/223	OIL TO UNIT 224			
CO ₂	44.010 lbmol/hr					
H ₂ S	34.076 lbmol/hr					
COS	60.070 lbmol/hr					
C ₂ H ₄	28.054 lbmol/hr					
CO	28.010 lbmol/hr					
H ₂	2.016 lbmol/hr					
CH ₄	16.043 lbmol/hr					
C ₂ H ₆	30.070 lbmol/hr					
N ₂ +INERTS	28.013 lbmol/hr					
O ₂	32.000 lbmol/hr					
TOTAL DRY FLOW	lbmol/hr					
H ₂ O	18.016 lbmol/hr					
METHANOL	32.042 lbmol/hr					
TOTAL WET GAS	lbmol/hr					
TOTAL DRY GAS	lb/hr					
H ₂ O	lb/hr					
METHANOL	lb/hr					
TAR	lb/hr					
OIL	lb/hr	46 353	5 358			
NAPHTHA	lb/hr					
CRUDE PHENOLS	lb/hr					
ANHYDR. AMMONIA	lb/hr					
SULPHUR	lb/hr					
OTHERS	lb/hr					
MINERALS	lb/hr					
COAL	lb/hr					
SULPHUR ORGANIC	lb/hr					
NITROGEN ORGANIC	lb/hr					
FATTY ACIDS	lb/hr					
ACETONE	lb/hr					
C ₆ +	lb/hr					
TOTAL	lb/hr	46 353	5 358			

MATERIAL BALANCE

BASE CASE II

STREAM N°			62	63	64	65	66
			UNIT 224 STACKGAS	UNIT 223/222 STACKGAS	SO2 FROM UNIT 226	STACKGAS TO ATMOSPHERE	N2 TO ATMOSPHERE
	CO ₂	44.010 lbmol/hr	358.85	61 895.55		62 254.46	
	SO ₂	64.046 lbmol/hr	0.65	54.75	40.28	15.12	
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr					
	CH ₄	16.043 lbmol/hr					
	C ₂ H ₆	30.070 lbmol/hr					
	N ₂ + INERTS	28.013 lbmol/hr	2 010.60	127 685.79		129 696.39	55 553.58
	O ₂	32.000 lbmol/hr	69.68	2 157.04		2 226.72	781.28
	TOTAL DRY FLOW (1)	lbmol/hr	2 439.78	191 793.13		194 192.69	56 334.86
	H ₂ O	18.016 lbmol/hr	291.45	27 332.67		27 624.12	641.87
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr	2 731.23	219 125.80		221 816.81	56 976.73
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr					
	nC ₄ H ₁₀	58.12 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr					
	nC ₅ H ₁₂	72.15 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr					
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	81 432	6 865 883	2 577	6 942 944	1 592 787
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

MATERIAL BALANCE

BASE CASE II

STREAM N°			67	68	69	70	71
			Air	Air to Units 222/223	Stackgas	Air to Unit 221	Water from Unit 221
	CO ₂	44.010 lbmol/hr			2360.01		
	H ₂ S	34.076 lbmol/hr			0.01		
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr			0.50		
	CO	28.010 lbmol/hr			0.04		
	H ₂	2.016 lbmol/hr			29.49		
	CH ₄	16.043 lbmol/hr			54.10		
	C ₂ H ₆	30.070 lbmol/hr			1.48		
	N ₂ +INERTS	28.013 lbmol/hr	185 432.79	122 357.89	5 234.76	55 831.48	
	O ₂	32.000 lbmol/hr	49 292.26	32 525.51	176.67	14 841.28	
	TOTAL DRY FLOW (1)	lbmol/hr	234 725.05	154 883.40	7 857.77	70 672.76	
	H ₂ O	18.016 lbmol/hr	5 840.63	3 853.91	1 353.12	1 758.61	1 116.73
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr	240 565.68	158 737.31	9 210.89	72 431.37	
	C ₃ H ₆	42.08 lbmol/hr			0.50		
	C ₃ H ₈	44.09 lbmol/hr			0.07		
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr					
	nC ₄ H ₁₀	58.12 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr					
	nC ₅ H ₁₂	72.15 lbmol/hr			0.14		
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr					
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	6877 107	4 537 860	281 560	2070 606	20 119
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

MATERIAL BALANCE

BASE CASE II

STREAM N ^o			72	73			
			Air to Unit 235	Air to Unit 214			
	CO ₂	44.010 lbmol/hr					
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr					
	CH ₄	16.043 lbmol/hr					
	C ₂ H ₆	30.070 lbmol/hr					
	N ₂ +INERTS	28.013 lbmol/hr		93.11			
	O ₂	32.000 lbmol/hr		24.75			
	TOTAL DRY FLOW (1)	lbmol/hr		117.86			
	H ₂ O	18.016 lbmol/hr		2.93			
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr		120.79			
	C ₃ H ₈	42.08 lbmol/hr					
	C ₃ H ₆	44.09 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₆	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr					
	nC ₄ H ₁₀	58.12 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	70.13 lbmol/hr					
	iC ₅ H ₁₄	72.15 lbmol/hr					
	nC ₅ H ₁₂	72.15 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	84.16 lbmol/hr					
	iC ₆ H ₁₆	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr					
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	440 000	3 453			
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

BASE CASE II

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MATERIAL BALANCE

BASE CASE II

STREAM N°			78	79	80	81
			Steam to Unit 208	Unit 251 Purge Gas	Treated Water	BFW to Unit 206
	CO ₂	44.010 lbmol/hr				
	H ₂ S	34.076 lbmol/hr		0.01		
	COS	60.070 lbmol/hr				
	C ₂ H ₄	28.054 lbmol/hr				
	CO	28.010 lbmol/hr				
	H ₂	2.016 lbmol/hr		26.79		
	CH ₄	16.043 lbmol/hr		5.92		
	C ₂ H ₆	30.070 lbmol/hr		0.28		
	N ₂ +INERTS	28.013 lbmol/hr				
	O ₂	32.000 lbmol/hr				
	TOTAL DRY FLOW (1)	lbmol/hr		33.25		
	H ₂ O	18.016 lbmol/hr	344.14	0.05	108 900.25	2 130.88
	METHANOL	32.042 lbmol/hr				
	TOTAL WET FLOW (1)	lbmol/hr		33.30		
	C ₃ H ₆	42.08 lbmol/hr				
	C ₃ H ₈	44.09 lbmol/hr		0.02		
	iC ₄ H ₈	56.10 lbmol/hr				
	iC ₄ H ₈	56.10 lbmol/hr				
	iC ₄ H ₁₀	58.12 lbmol/hr				
	nC ₄ H ₁₀	58.12 lbmol/hr				
	iC ₅ H ₁₀	70.13 lbmol/hr				
	iC ₅ H ₁₀	70.13 lbmol/hr				
	iC ₅ H ₁₂	72.15 lbmol/hr				
	nC ₅ H ₁₂	72.15 lbmol/hr		0.14		
	iC ₆ H ₁₂	84.16 lbmol/hr				
	iC ₆ H ₁₂	84.16 lbmol/hr				
	iC ₆ H ₁₄	86.17 lbmol/hr				
	nC ₆ H ₁₄	86.17 lbmol/hr				
	C ₆ +	(2) lbmol/hr				
	C ₇ +	(3) lbmol/hr				
	ALCOHOLS	46.61 lbmol/hr				
	TOTAL	lb/hr	6 200	169	1 961 947	38 390
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					
	(3) MW OF C ₇ +					

MATERIAL BALANCE

BASE CASE II

STREAM N°		82	84	85.1	85.2	86
		Water from Unit 213	Air to Unit 224	Air to Unit 206	Air from Unit 206	RFW Make Up Water
	CO ₂ 44.010 lbmol/hr					
	H ₂ S 34.076 lbmol/hr					
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr			air surplus	Air surplus	
	CO 28.010 lbmol/hr			+ N ₂	+ N ₂	
	H ₂ 2.016 lbmol/hr			not	not	
	CH ₄ 16.043 lbmol/hr			balanced	balanced	
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERTS 28.013 lbmol/hr		2 009.71			
	O ₂ 32.000 lbmol/hr		534.23	78.88		
	TOTAL DRY FLOW (1) lbmol/hr		2 543.94			
	H ₂ O 18.016 lbmol/hr	148.31	63.30			118 805.51
	METHANOL 32.042 lbmol/hr					
	TOTAL WET FLOW (1) lbmol/hr		2 607.24			
	C ₃ H ₆ 42.08 lbmol/hr					
	C ₃ H ₈ 44.09 lbmol/hr					
	iC ₄ H ₈ 56.10 lbmol/hr					
	iC ₄ H ₈ 56.10 lbmol/hr					
	iC ₄ H ₁₀ 58.12 lbmol/hr					
	nC ₄ H ₁₀ 58.12 lbmol/hr					
	iC ₅ H ₁₀ 70.13 lbmol/hr					
	iC ₅ H ₁₀ 70.13 lbmol/hr					
	iC ₅ H ₁₂ 72.15 lbmol/hr					
	nC ₅ H ₁₂ 72.15 lbmol/hr					
	iC ₆ H ₁₂ 84.16 lbmol/hr					
	iC ₆ H ₁₂ 84.16 lbmol/hr					
	iC ₆ H ₁₄ 86.17 lbmol/hr					
	nC ₆ H ₁₄ 86.17 lbmol/hr					
	C ₆ + (2) lbmol/hr					
	C ₇ + (3) lbmol/hr					
	ALCOHOLS 46.61 lbmol/hr					
	TOTAL lb/hr	2 672	76 212	2 524	0	2 140 400
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					
	(3) MW OF C ₇ +					

MATERIAL BALANCE

BASE CASE II

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STREAM №		87	88	89	90	91
		BLOW DOWN FROM UNIT 222	C.W. MAKE UP WATER	EVAPORATION LOSSES OF U233	WASTE AIR UNIT 235	C.W. BLOW DOWN
	CO ₂ 44.010 lbmol/hr				313.89	
	SO ₂ 64.046 lbmol/hr				1.88	
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ + INERTS 28.013 lbmol/hr				11 632.83	
	O ₂ 32.000 lbmol/hr				2 703.65	
	TOTAL DRY GAS lbmol/hr				14 652.25	
	H ₂ O 18.016 lbmol/hr	8 603.46	65 480.68	133 936.50	776.05	37 907.80
	METHANOL 32.042 lbmol/hr					
	TOTAL WET GAS lbmol/hr				15 428.30	
	TOTAL DRY GAS lb/hr				426 322	
	H ₂ O lb/hr	155 000	1179 700	2413 000	13 981	682 947
	METHANOL lb/hr					
	TAR lb/hr					
	OIL lb/hr					
	NAPHTHA lb/hr					
	CRUDE PHENOLS lb/hr				Air surplus not balanced	
	ANHYDR. AMMONIA lb/hr					
	SULPHUR lb/hr					
	OTHERS lb/hr					
	MINERALS lb/hr					
	COAL lb/hr					
	SULPHUR ORGANIC lb/hr					
	NITROGEN ORGANIC lb/hr					
	FATTY ACIDS lb/hr					
	ACETONE lb/hr					
	C ₆ + lb/hr					
	TOTAL lb/hr	155 000	1179 700	2413 000	440 303	682 947

MATERIAL BALANCE

BASE CASE II

STREAM N°		92	93	94	95	96
			Sluice Vent	Steam to Unit 251	Deaerator losses	UNIT 232 Blow Down
	CO ₂ 44.010 lbmol/hr					
	H ₂ S 34.076 lbmol/hr					
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr		Air not			
	N ₂ +INERTS 28.013 lbmol/hr					
	O ₂ 32.000 lbmol/hr		balanced			
	TOTAL DRY FLOW (1) lbmol/hr					
	H ₂ O 18.016 lbmol/hr	2 525.53	1 415.41	394.09	555.06	962.87
	METHANOL 32.042 lbmol/hr					
	TOTAL WET FLOW (1) lbmol/hr					
	C ₃ H ₆ 42.08 lbmol/hr					
	C ₃ H ₈ 44.09 lbmol/hr					
	iC ₄ H ₈ 56.10 lbmol/hr					
	iC ₄ H ₈ 56.10 lbmol/hr					
	iC ₄ H ₁₀ 58.12 lbmol/hr					
	nC ₄ H ₁₀ 58.12 lbmol/hr					
	iC ₅ H ₁₀ 70.13 lbmol/hr					
	iC ₅ H ₁₀ 70.13 lbmol/hr					
	iC ₅ H ₁₂ 72.15 lbmol/hr					
	nC ₅ H ₁₂ 72.15 lbmol/hr					
	iC ₆ H ₁₂ 84.16 lbmol/hr					
	iC ₆ H ₁₂ 84.16 lbmol/hr					
	iC ₆ H ₁₄ 86.17 lbmol/hr					
	nC ₆ H ₁₄ 86.17 lbmol/hr					
	C ₆ + (2) lbmol/hr					
	C ₇ + (3) lbmol/hr					
	ALCOHOLS 46.61 lbmol/hr					
	TOTAL lb/hr	45 500	25 500	7 100	10 000	17 347
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					
	(3) MW OF C ₇ +					

MATERIAL BALANCE

BASE CASE II

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STREAM N°			97				
			UNIT 235 STRIPPINGSTEAM				
	CO ₂	44.010 lbmol/hr					
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr					
	CH ₄	16.043 lbmol/hr					
	C ₂ H ₆	30.070 lbmol/hr					
	N ₂ +INERTS	28.013 lbmol/hr					
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr					
	H ₂ O	18.016 lbmol/hr	499.56				
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr					
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr					
	nC ₄ H ₁₀	58.12 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr					
	nC ₅ H ₁₂	72.15 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr					
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	9 000				
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

MATERIAL BALANCE

BASE CASE II

STREAM N°		100	101	102	103	104
		NAPHTHA FROM UNIT 251	WASH WATER TO UNIT 251	FEED TO UNIT 210	LIGHT OIL UNIT 250	HEAVY OIL FROM UNIT 250
	CO ₂ 44.010 lbmol/hr			3 084.66	16.87	0.35
	H ₂ S 34.076 lbmol/hr					
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr			378.13	2.41	0.06
	CO 28.010 lbmol/hr			483.04	0.29	0.03
	H ₂ 2.016 lbmol/hr			9 334.26	2.71	0.30
	CH ₄ 16.043 lbmol/hr			14 227.78	29.52	0.90
	C ₂ H ₆ 30.070 lbmol/hr			810.76	6.81	0.13
	N ₂ +INERTS 28.013 lbmol/hr			308.79	0.19	0.02
	O ₂ 32.000 lbmol/hr					
	TOTAL DRY FLOW (1) lbmol/hr	160.41		30 292.01	730.70	35.75
	H ₂ O 18.016 lbmol/hr		48.29	117.29		
	METHANOL 32.042 lbmol/hr					
	TOTAL WET FLOW (1) lbmol/hr			30 409.30		
	C ₃ H ₆ 42.08 lbmol/hr			728.15	16.95	0.20
	C ₃ H ₈ 44.09 lbmol/hr			121.20	3.21	0.04
	iC ₄ H ₈ 56.10 lbmol/hr			371.27	28.84	0.19
	iC ₄ H ₈ 56.10 lbmol/hr					
	iC ₄ H ₁₀ 58.12 lbmol/hr			2.43	0.17	
	nC ₄ H ₁₀ 58.12 lbmol/hr	1.98		45.29	4.01	0.03
	iC ₅ H ₁₀ 70.13 lbmol/hr			187.20	43.85	0.17
	iC ₅ H ₁₀ 70.13 lbmol/hr					
	iC ₅ H ₁₂ 72.15 lbmol/hr			8.06	1.83	0.01
	nC ₅ H ₁₂ 72.15 lbmol/hr			23.09	6.60	0.02
	iC ₆ H ₁₂ 84.16 lbmol/hr			91.42	60.95	0.14
	iC ₆ H ₁₂ 84.16 lbmol/hr					
	iC ₆ H ₁₄ 86.17 lbmol/hr			6.66	4.50	0.01
	nC ₆ H ₁₄ 86.17 lbmol/hr			13.92	12.14	0.02
	C ₆ + (2) lbmol/hr					
	C ₇ + (3) lbmol/hr			65.90	488.85	33.13
	ALCOHOLS 46.61 lbmol/hr					
	NAPHTHA 97.30	158.43				
	TOTAL lb/hr	15 541	870	533 527	84 738	11 399
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					
	(3) MW OF C ₇ +			106.10	143.74	341.39

BASE CASE II

[illegible]

MATERIAL BALANCE

BASE CASE II

STREAM N°			106	107	108	109	
			LIGHT GAS FR. UNIT 252	FEED TO UNIT 257	FEED TO UNIT 253	VACUUM RESIDU- AL TO UNIT 222	
	CO ₂	44.010 lbmol/hr	1 087.22				
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr	196.57				
	CO	28.010 lbmol/hr	9.32				
	H ₂	2.016 lbmol/hr	52.01				
	CH ₄	16.043 lbmol/hr	1 040.42				
	C ₂ H ₆	30.070 lbmol/hr	466.73	1.00			
	N ₂ +INERTS	28.013 lbmol/hr	5.21				
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	3 076.23	1 093.06	1 100.41	4.23	
	H ₂ O	18.016 lbmol/hr	116.86	0.42			
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr	3 193.09	1 093.48			
	C ₃ H ₆	42.08 lbmol/hr	198.57	546.73			
	C ₃ H ₈	44.09 lbmol/hr	19.13	105.32			
	iC ₄ H ₈	56.10 lbmol/hr	0.25	384.64	15.41		
	iC ₄ H ₁₀	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr	0.01	2.52	0.07		
	nC ₄ H ₁₀	58.12 lbmol/hr	0.03	44.29	5.01		
	iC ₅ H ₁₀	70.13 lbmol/hr	0.04	7.55	223.63		
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr		0.75	9.15		
	nC ₅ H ₁₂	72.15 lbmol/hr	0.01	0.26	29.44		
	iC ₆ H ₁₂	84.16 lbmol/hr	0.01		152.50		
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr	0.01		11.16		
	nC ₆ H ₁₄	86.17 lbmol/hr	0.01				
	C ₆ +	(2) lbmol/hr			26.07		
	C ₇ +	(3) lbmol/hr	0.68		577.97	4.23	
	ALCOHOLS	51.24 lbmol/hr			50		
	TOTAL	lb/hr	96 028	52 589	123 922	2 019	
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +		153.24		148.29	477.30	

MATERIAL BALANCE

BASE CASE II

STREAM N°			110	111	112	113	114
			H ₂ TO UNIT 253	H ₂ to UNIT 259	FEED TO UNIT 254	WASTE WATER FROM UNIT 253	STRIPPING STEAM TO UNIT 254
	CO ₂	44.010 lbmol/hr	3.15	2.55	0.60		
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₆	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr	1 444.55	362.12	44.19		
	CH ₄	16.043 lbmol/hr	56.50	37.32	24.66		
	C ₂ H ₆	30.070 lbmol/hr	13.50	18.87	49.56		
	N ₂ +INERTS	28.013 lbmol/hr					
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	1 518.53	427.73	1 204.68		
	H ₂ O	18.016 lbmol/hr		0.80	7.75	135.10	62.17
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr		428.53	1 212.43	135.10	62.17
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr	0.62	0.91	7.37		
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	56.10 lbmol/hr					
	iC ₄ H ₁₂	58.12 lbmol/hr	0.03	0.23	4.37		
	nC ₄ H ₁₀	58.12 lbmol/hr		0.93	24.92		
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	70.13 lbmol/hr					
	iC ₅ H ₁₄	72.15 lbmol/hr	0.14	0.74	48.14		
	nC ₅ H ₁₂	72.15 lbmol/hr		2.84	235.13		
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	84.16 lbmol/hr					
	iC ₆ H ₁₆	86.17 lbmol/hr	0.04	0.32	61.51		
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr		0.61	150.67		
	C ₇ +	(3) lbmol/hr		0.29	553.56		
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	4 411	2 500	123 398	2 434	1 120
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +			103.00	145.03		

MATERIAL BALANCE

BASE CASE II

STREAM N°			115	116	117	118	119
			FEED TO UNIT 260	WASTE WATER FR. UNIT 254	DISTILLATE FUELOIL PROD.	HEAVY FUELOIL PRODUCT	FEED TO UNIT 260
	CO ₂	44.010 lbmol/hr	0.52				0.08
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr	43.62				0.57
	CH ₄	16.043 lbmol/hr	22.39				2.27
	C ₂ H ₆	30.070 lbmol/hr	34.85				14.71
	N ₂ +INERTS	28.013 lbmol/hr					
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	134.48		129.79	22.81	44.21
	H ₂ O	18.016 lbmol/hr	5.27	62.17	1.98	0.50	
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr	139.75	62.17	131.77	23.31	
	C ₃ H ₈	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr	3.08				4.29
	iC ₄ H ₁₀	56.10 lbmol/hr					
	iC ₄ H ₁₀	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr	0.97				3.34
	nC ₄ H ₁₀	58.12 lbmol/hr	4.31				18.26
	iC ₅ H ₁₂	70.13 lbmol/hr					
	iC ₅ H ₁₂	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr	3.73				0.41
	nC ₅ H ₁₂	72.15 lbmol/hr	14.55		0.02		0.28
	iC ₆ H ₁₄	84.16 lbmol/hr					
	iC ₆ H ₁₄	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr	1.69		0.02		
	nC ₆ H ₁₄	86.17 lbmol/hr			0.05		
	C ₆ +	(2) lbmol/hr	3.18				
	C ₇ +	(3) lbmol/hr	1.59		129.70	24.32	
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	3 959	1 120	25 425	7 445	1 978
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +		103.82		195.70	326.00	

MATERIAL BALANCE

BASE CASE II

STREAM N°			120	121	122	123	124
			FEED TO UNIT 256	FEED TO UNIT 255	FEED TO UNIT 261	UNIT 255 OFF GAS	FEED TO UNIT 260
	CO ₂	44.010 lbmol/hr					
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr			373.07	2.04	0.64
	CH ₄	16.043 lbmol/hr			36.80	0.75	1.21
	C ₂ H ₆	30.070 lbmol/hr			42.38	1.35	7.95
	N ₂ + INERTS	28.013 lbmol/hr					
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	474.45	397.43	552.35	7.86	103.43
	H ₂ O	18.016 lbmol/hr					
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr					
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr			54.00	2.22	35.82
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr	0.06		18.48	0.91	31.17
	nC ₄ H ₁₀	58.12 lbmol/hr	2.35		12.58	0.59	26.64
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr	44.00		7.28		
	nC ₅ H ₁₂	72.15 lbmol/hr	220.28		3.56		
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr	59.77	0.03			
	nC ₆ H ₁₄	86.17 lbmol/hr	146.74	0.70			
	C ₆ +	(2) lbmol/hr			4.20		
	C ₇ +	(3) lbmol/hr	1.25	396.70			
		lbmol/hr					
	TOTAL	lb/hr	37 128	46 972	7 981	242	5 199
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +				94.00		
	(3) MW OF C ₇ +		100.20	116.22			

MATERIAL BALANCE

BASE CASE II

STREAM N°			125	126	127	128	129
			REFORMATE FR. UNIT 255	H ₂ TO UNIT 256	ISOMERATE FR. UNIT 256	ISOMERATE FR. UNIT 256	H ₂ RICH OFF FR. UNIT 256
	CO ₂	44.010 lbmol/hr		0.03			0.06
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr		18.98			3.89
	CH ₄	16.043 lbmol/hr		0.76			0.02
	C ₂ H ₆	30.070 lbmol/hr		0.16			0.02
	N ₂ +INERTS	28.013 lbmol/hr					
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	349.87	19.95	237.34	213.86	4.17
	H ₂ O	18.016 lbmol/hr					
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr					
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr		0.02			0.02
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr	0.18		0.70		0.01
	nC ₄ H ₁₀	58.12 lbmol/hr	2.65		3.29		
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr	42.74		225.59	6.65	0.14
	nC ₅ H ₁₂	72.15 lbmol/hr	26.69		7.76	2.83	
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr				173.67	0.04
	nC ₆ H ₁₄	86.17 lbmol/hr				30.71	
	C₆+	(2) lbmol/hr	277.61				
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	33 551	58	17 068	18 295	25
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +		102.22				
	(3) MW OF C ₇ +						

MATERIAL BALANCE

BASE CASE II

STREAM N°			130	131			
			FEED TO UNIT 261	FEED TO UNIT 260			
	CO ₂	44.010 lbmol/hr					
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr	9.48	0.42			
	CH ₄	16.043 lbmol/hr	0.63	0.11			
	C ₂ H ₆	30.070 lbmol/hr	0.73	0.57			
	N ₂ + INERTS	28.013 lbmol/hr					
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	13.61	24.22			
	H ₂ O	18.016 lbmol/hr					
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr					
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr	1.10	2.54			
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr	0.43	2.15			
	nC ₄ H ₁₀	58.12 lbmol/hr	0.09	0.59			
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr	1.12	17.05			
	nC ₅ H ₁₂	72.15 lbmol/hr	0.02	0.44			
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr	0.01	0.35			
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr					
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	213	1 583			
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

MATERIAL BALANCE

BASE CASE II

STREAM N°			132	133	134	135	136
			FEED TO UNIT 258	FUEL GAS FR. UNIT 252	LT.CAT.POLY GASOLINE	FEED TO UNIT 259	FEED TO UNIT 261
	CO ₂	44.010 lbmol/hr		46.97			2.55
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr		8.49			
	CO	28.010 lbmol/hr		0.40			
	H ₂	2.016 lbmol/hr		2.25			227.12
	CH ₄	16.043 lbmol/hr		44.95			37.32
	C ₂ H ₆	30.070 lbmol/hr	1.00	20.16			18.87
	N ₂ + INERTS	28.013 lbmol/hr		0.23			
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	210.05	132.91	241.45	135.00	292.73
	H ₂ O	18.016 lbmol/hr		5.05			0.80
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr		137.96			293.53
	C ₃ H ₆	42.08 lbmol/hr	54.67	8.58			
	C ₃ H ₈	44.09 lbmol/hr	105.32	0.83			0.91
	iC ₄ H ₈	56.10 lbmol/hr	19.26	0.01			
	iC ₄ H ₁₀	58.12 lbmol/hr	2.15	0.01	0.37		0.23
	nC ₄ H ₁₀	58.12 lbmol/hr	27.65		16.64		0.93
	iC ₅ H ₁₀	70.13 lbmol/hr			7.55		
	iC ₅ H ₁₂	72.15 lbmol/hr			0.75		0.74
	nC ₅ H ₁₂	72.15 lbmol/hr			0.26		2.84
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr					1.22
	C ₇ +	(3) lbmol/hr		0.03			
	ALCOHOLS	46.61 lbmol/hr					
	CAT. POLY GASOLINE	lbmol/hr			215.88	135.00	
	TOTAL	lb/hr	9 787	4 149	22 952	19 482	2 227
	(4) MW OF CAT. POLY GAS.				98.95	146.98	
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +			153.24			90.17

MATERIAL BALANCE

BASE CASE II

STREAM N ^o			137	138	139	140	141
			HVY POLY GAS FROM UNIT 259	PROPANE LPG PRODUCT	NORMAL BUTANE PRODUCT	NORMAL BUTANE TO UNIT 270	FUEL GAS FROM UNIT 160
	CO ₂	44.010 lbmol/hr					0.82
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr					94.02
	CH ₄	16.043 lbmol/hr					56.11
	C ₂ H ₆	30.070 lbmol/hr		1.06			106.36
	N ₂ + INERTS	28.013 lbmol/hr					
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	135	186.89	21.48	65.10	323.02
	H ₂ O	18.016 lbmol/hr					5.85
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr					328.87
	C ₃ H ₆	42.08 lbmol/hr		17.71			
	C ₃ H ₈	44.09 lbmol/hr		168.12			43.83
	iC ₄ H ₈	56.10 lbmol/hr			1.55	4.68	
	iC ₄ H ₁₀	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr					9.32
	nC ₄ H ₁₀	58.12 lbmol/hr			19.93	60.42	6.63
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr					2.38
	nC ₅ H ₁₂	72.15 lbmol/hr					3.10
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr					0.01
	nC ₆ H ₁₄	86.17 lbmol/hr					0.01
	C ₆ +	(2) lbmol/hr					0.43
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	CAT. POLY GASOLINE	lbmol/hr	135				
	TOTAL	lb/hr	20 115	8 190	1 245	3 774	7 725
	(4) MW OF CAT. POLY GAS.		149				
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

MATERIAL BALANCE

BASE CASE II

STREAM N ^o			142	143	144	145	146
			ALKYLATE FROM UNIT 258	ISOBUTANE FROM UNIT 260	LT. GASOLINE FROM UNIT 260	FEED TO UNIT 260	HYDROGEN FROM UNIT 261
	CO ₂	44.010 lbmol/hr				0.22	
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr				48.77	560.90
	CH ₄	16.043 lbmol/hr				30.13	44.62
	C ₂ H ₆	30.070 lbmol/hr				48.34	13.64
	N ₂ + INERTS	28.013 lbmol/hr		0.06			
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	55.33	166.11	53.77	236.56	619.80
	H ₂ O	18.016 lbmol/hr				0.58	
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr				237.14	
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr		57.29		55.39	0.62
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	56.10 lbmol/hr					
	iC ₄ H ₁₂	58.12 lbmol/hr		47.36	0.07	19.12	0.02
	nC ₄ H ₁₀	58.12 lbmol/hr	2.05	54.75	2.02	13.60	
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	70.13 lbmol/hr					
	iC ₅ H ₁₄	72.15 lbmol/hr	9.99	6.04	21.91	9.14	
	nC ₅ H ₁₂	72.15 lbmol/hr	0.60	0.60	17.99	6.42	
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	84.16 lbmol/hr					
	iC ₆ H ₁₆	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr	0.01	0.01			
	C ₆ +	(2) lbmol/hr			11.78	5.43	
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	ALKYLATE	108.60 lbmol/hr	42.68				
	TOTAL	lb/hr	5 519	8 942	4 078	8 028	2 287
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

MATERIAL BALANCE

BASE CASE II

[illegible]

MATERIAL BALANCE

BASE CASE II

[illegible]

BASE CASE II

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MATERIAL BALANCE

BASE CASE II

[illegible]

MATERIAL BALANCE

BASE CASE II

STREAM №			171	172	173.1	173.2	174
			FUEL TO UNIT 251	FUEL TO UNIT 252	FUEL TO UNIT 253	FUEL TO UNIT 253	FUEL TO UNIT 254
	CO ₂	44.010 lbmol/hr	0.27	27.45	2.67		8.49
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr	0.05	4.87	0.48		1.51
	CO	28.010 lbmol/hr		0.24	0.02		0.07
	H ₂	2.016 lbmol/hr	0.57	56.49	5.49		17.46
	CH ₄	16.043 lbmol/hr	0.58	58.49	5.70		18.08
	C ₂ H ₆	30.070 lbmol/hr	0.74	73.46	7.15		22.71
	N ₂ +INERTS	28.013 lbmol/hr		0.14	0.01		0.04
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	2.66	266.49	25.93	1.51	82.38
	H ₂ O	18.016 lbmol/hr	0.06	6.26	0.61		1.94
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr	2.72	272.75	26.54		84.32
	C ₃ H ₆	42.08 lbmol/hr	0.05	4.93	0.48		1.52
	C ₃ H ₈	44.09 lbmol/hr	0.27	26.92	2.62		8.33
	iC ₄ H ₈	56.10 lbmol/hr		0.01			
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr	0.06	5.88	0.58		1.82
	nC ₄ H ₁₀	58.12 lbmol/hr	0.04	4.16	0.40		1.28
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr	0.01	1.36	0.13		0.43
	nC ₅ H ₁₂	72.15 lbmol/hr	0.02	1.80	0.17		0.56
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr		0.26			
	-C ₆ +	(2) lbmol/hr			0.03		0.08
	C ₇ +	(3) lbmol/hr		0.03		1.51	
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	69	6 963	678	492	2 153
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +			153.24		326	

MATERIAL BALANCE

BASE CASE II

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STREAM N°			175	176	177	178	179
			Fuel to Unit 255	Fuel to Unit 256	Fuel to Unit 259	Fuel to Unit 271	Air to Unit 251
	CO ₂	44.010 lbmol/hr	4.48	0.48	0.32	2.72	
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr	0.79	0.09	0.06	0.48	
	CO	28.010 lbmol/hr	0.04			0.02	
	H ₂	2.016 lbmol/hr	9.22	0.99	0.66	5.60	
	CH ₄	16.043 lbmol/hr	9.56	1.02	0.69	5.80	
	C ₂ H ₆	30.070 lbmol/hr	12.00	1.28	0.85	7.29	
	N ₂ +INERTS	28.013 lbmol/hr	0.03	0.01		0.01	30.02
	O ₂	32.000 lbmol/hr					7.98
	TOTAL DRY FLOW (1)	lbmol/hr	43.54	4.66	3.11	26.42	38.00
	H ₂ O	18.016 lbmol/hr	1.02	0.11	0.07	0.62	0.95
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr	44.56	4.77	3.18	27.04	38.95
	C ₃ H ₆	42.08 lbmol/hr	0.81	0.09	0.06	0.49	
	C ₃ H ₈	44.09 lbmol/hr	4.40	0.47	0.32	2.67	
	1C ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr	0.97	0.10	0.07	0.58	
	nC ₄ H ₁₀	58.12 lbmol/hr	0.67	0.07	0.05	0.41	
	1C ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	1C ₅ H ₁₂	72.15 lbmol/hr	0.23	0.03	0.01	0.14	
	nC ₅ H ₁₂	72.15 lbmol/hr	0.30	0.03	0.02	0.18	
	1C ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	1C ₆ H ₁₄	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ + (2)	lbmol/hr	0.04			0.03	
	C ₇ + (3)	lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	1137	122	81	690	1 113
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

MATERIAL BALANCE

BASE CASE II

STREAM N°			180	181	182	183	184
			AIR TO UNIT 252	AIR TO UNIT 253	AIR TO UNIT 254	AIR TO UNIT 255	AIR TO UNIT 256
	CO ₂	44.010 lbmol/hr					
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr					
	CH ₄	16.043 lbmol/hr					
	C ₂ H ₆	30.070 lbmol/hr					
	N ₂ + INERTS	28.013 lbmol/hr	2 865.59	510.08	892.66	471.88	50.42
	O ₂	32.000 lbmol/hr	761.74	135.59	237.29	125.44	13.40
	TOTAL DRY FLOW (1)	lbmol/hr	3 627.33	645.67	1 129.95	597.32	63.82
	H ₂ O	18.016 lbmol/hr	90.26	16.06	28.11	14.86	1.59
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr	3 717.59	661.73	1 158.06	612.18	65.41
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr					
	nC ₄ H ₁₀	58.12 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr					
	nC ₅ H ₁₂	72.15 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr					
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	106 276.	18 917	33 106	17 500	1 870
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

MATERIAL BALANCE

BASE CASE II

STREAM N°		190	191	192	193	194
		STACKGAS FROM 254	STACKGAS FROM UNIT 255	STACKGAS FROM UNIT 256	STACKGAS FROM UNIT 259	STACKGAS FROM UNIT 271
	CO ₂ 44.010 lbmol/hr	122.54	64.74	6.90	4.60	39.28
	H ₂ S 34.076 lbmol/hr					
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr					
	CO 28.010 lbmol/hr					
	H ₂ 2.016 lbmol/hr					
	CH ₄ 16.043 lbmol/hr					
	C ₂ H ₆ 30.070 lbmol/hr					
	N ₂ +INERTS 28.013 lbmol/hr	892.70	471.92	50.43	33.56	286.40
	O ₂ 32.000 lbmol/hr	30.95	16.39	1.78	1.17	9.93
	TOTAL DRY FLOW (1) lbmol/hr	1 046.19	553.05	59.11	39.33	335.61
	H ₂ O 18.016 lbmol/hr	214.70	113.49	12.11	8.08	68.90
	METHANOL 32.042 lbmol/hr					
	TOTAL WET FLOW (1) lbmol/hr	1 260.89	666.54	71.22	47.41	404.51
	C ₃ H ₆ 42.08 lbmol/hr					
	C ₃ H ₈ 44.09 lbmol/hr					
	iC ₄ H ₈ 56.10 lbmol/hr					
	iC ₄ H ₈ 56.10 lbmol/hr					
	iC ₄ H ₁₀ 58.12 lbmol/hr					
	nC ₄ H ₁₀ 58.12 lbmol/hr					
	iC ₅ H ₁₀ 70.13 lbmol/hr					
	iC ₅ H ₁₀ 70.13 lbmol/hr					
	iC ₅ H ₁₂ 72.15 lbmol/hr					
	nC ₅ H ₁₂ 72.15 lbmol/hr					
	iC ₆ H ₁₂ 84.16 lbmol/hr					
	iC ₆ H ₁₂ 84.16 lbmol/hr					
	iC ₆ H ₁₄ 86.17 lbmol/hr					
	nC ₆ H ₁₄ 86.17 lbmol/hr					
	C ₆ + (2) lbmol/hr					
	C ₇ + (3) lbmol/hr					
	ALCOHOLS 46.61 lbmol/hr					
	TOTAL lb/hr	35 259	18 637	1 992	1 326	11 312
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					
	(3) MW OF C ₇ +					

MATERIAL BALANCE

BASE CASE II

STREAM N°			195	196	197		
			WATER TO UNIT 250	WATER TO UNIT 257	WATER FROM UNIT 257		
	CO ₂	44.010 lbmol/hr					
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr					
	H ₂	2.016 lbmol/hr					
	CH ₄	16.043 lbmol/hr					
	C ₂ H ₆	30.070 lbmol/hr					
	N ₂ +INERTS	28.013 lbmol/hr					
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr					
	H ₂ O	18.016 lbmol/hr	5 634.99	2 775.31	2 775.31		
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr	5 634.99	2 775.31	2 775.31		
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr					
	nC ₄ H ₁₀	58.12 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr					
	nC ₅ H ₁₂	72.15 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr					
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	101 520	50 000	50 000		
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +

APPENDIX C

MATERIAL BALANCE (PARTIAL) - SENSITIVITY CASES

MATERIAL BALANCE

SENSITIVITY CASE I-A

STREAM N°			37	37.1	82.1	37.2	49
			FEED TO UNIT 150	METHANOL PRODUCT	WASTE WATER FROM UNIT 115	OFFGAS FROM UNIT 115	SNG PRODUCT
	CO ₂	44.010 lbmol/hr	32.00				84.85
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr					
	CO	28.010 lbmol/hr	0.20				0.96
	H ₂	2.016 lbmol/hr	0.24				270.30
	CH ₄	16.043 lbmol/hr	8.00				
	C ₂ H ₆	30.070 lbmol/hr	2.00				15 404.38
	N ₂ +INERT	23.013 lbmol/hr	0.03				
	O ₂	32.000 lbmol/hr					308.22
	TOTAL DRY GAS (1)	lbmol/hr	42.47				16 068.71
	H ₂ O	18.016 lbmol/hr	1 382.40				1.41
	METHANOL	32.042 lbmol/hr	18 360.30				
	TOTAL WET GAS (1)	lbmol/hr	19 785.17				16 070
	TOTAL DRY GAS	lb/hr	1 604	1 369		235	260 073
	H ₂ O	lb/hr	24 904	580	24 324		
	METHANOL	lb/hr	588 301	587 956	245	100	
	C ₃ H ₆	42.08 lbmol/hr					
	C ₃ H ₈	44.09 lbmol/hr					
	iC ₄ H ₁₀	58.12 lbmol/hr					
	1C ₄ H ₈	56.10 lbmol/hr					
	nC ₄ H ₁₀	58.12 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr					
	1C ₅ H ₁₀	70.13 lbmol/hr					
	nC ₅ H ₁₂	72.15 lbmol/hr					
	C ₆ +	(2) lbmol/hr					
	NAPHTHA	lb/hr					
	TOTAL	lb/hr	614 809	589 905	24 569	335	260 100
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						

MATERIAL BALANCE

SENSITIVITY CASE I-B

-307-

STREAM N°			30	30.1	33	35	36
			FROM GAS PURIF UNIT 105	METH.SYNTHES. UNIT 110 FEED	H2 TO UNIT 151	EXPANSION GAS FROM UNIT 110	UNIT 105 SOL-VENT MAKE UP
CO ₂	44.010 lbmol/hr		1293.89	7495.69		560.3	
H ₂ S	34.076 lbmol/hr						
COS	60.070 lbmol/hr						
C ₂ H ₄	28.054 lbmol/hr		21	23		1	
CO	28.010 lbmol/hr		18500	30199.21		50.6	
H ₂	2.016 lbmol/hr		46985	86345.98	96.22	164.0	
CH ₄	16.043 lbmol/hr		12460	13729.7	1.37	574.1	
C ₂ H ₆	30.070 lbmol/hr		277	307		9.6	
N ₂ +INERT	28.013 lbmol/hr		309	5576		69.95	
O ₂	32.000 lbmol/hr						
TOTAL DRY GAS (1)	lbmol/hr		79845.89	143376.58	97.59	1429.55	
H ₂ O	18.016 lbmol/hr			160.46		3.3	1.6
METHANOL	32.042 lbmol/hr		1.31	1.31		245.21	20.7
TOTAL WET GAS (1)	lbmol/hr		79847.2	143838.35		1678.06	22.3
TOTAL DRY GAS	lb/hr		887321	1736181	216	37893	
H ₂ O	lb/hr			2891		59	29
METHANOL	lb/hr		42	42		7857	663
C ₃ H ₆	42.08 lbmol/hr						
C ₃ H ₈	44.09 lbmol/hr						
iC ₄ H ₁₀	58.12 lbmol/hr						
1C ₄ H ₈	56.10 lbmol/hr						
nC ₄ H ₁₀	58.12 lbmol/hr						
iC ₅ H ₁₂	72.15 lbmol/hr						
1C ₅ H ₁₀	70.13 lbmol/hr						
nC ₅ H ₁₂	72.15 lbmol/hr						
C ₆ +	(2) lbmol/hr						
NAPHTHA			lb/hr				
TOTAL			lb/hr	887363	1739114	216	45810
(1) C ₃ + INCLUDED							
(2) MW OF C ₆ +							

MATERIAL BALANCE

SENSITIVITY CASE I-B

-308-

STREAM N°			37	38	39	-	42
			FEED TO UNIT 150	PURGE GAS FROM UNIT 110	PURGE GAS TO INCINERATION	FEED TO UNIT 116	REFORMED GAS
CO ₂	44.010 lbmol/hr		54	2949.7	250.7	3263.6	6456.11
H ₂ S	34.076 lbmol/hr						
COS	60.070 lbmol/hr						
C ₂ H ₄	28.054 lbmol/hr			22	1.9	29.8	2
CO	28.010 lbmol/hr		0.3	2116	179.9	2207.3	11699.21
H ₂	2.016 lbmol/hr		0.4	17558.98	1492.5	16143.46	39360.98
CH ₄	16.043 lbmol/hr		13.5	13142.1	1117.8	12662.83	1269.7
C ₂ H ₆	30.070 lbmol/hr		3.4	294	25.0	336	30
N ₂ + INERT	28.013 lbmol/hr		0.05	5506	468	5108	5267
O ₂	32.000 lbmol/hr			41588.78			
TOTAL DRY GAS (1)	lbmol/hr		71.65		3535.8	39750.99	64085.0
H ₂ O	18.016 lbmol/hr		4178.1	3	0.26	6.04	160.46
METHANOL	32.042 lbmol/hr		32218.31	119	10.12	354.09	
TOTAL WET GAS (1)	lbmol/hr		36468.06	41710.78	3546.18	40111.12	64245.46
TOTAL DRY GAS	lb/hr		2706	599020	50929	595182	860052
H ₂ O	lb/hr		204727	54	5	109	2891
METHANOL	lb/hr		1032339	3813	324	11346	
C ₃ H ₈	42.08 lbmol/hr						
C ₃ H ₆	44.09 lbmol/hr						
iC ₄ H ₁₀	58.12 lbmol/hr						
1C ₄ H ₈	56.10 lbmol/hr						
nC ₄ H ₁₀	58.12 lbmol/hr						
iC ₅ H ₁₂	72.15 lbmol/hr						
1C ₅ H ₁₀	70.13 lbmol/hr						
nC ₅ H ₁₂	72.15 lbmol/hr						
C ₆ +	(2) lbmol/hr						
NAPHTHA			lb/hr				
TOTAL			lb/hr	110318	602887	51258	606637
(1) C ₃ + INCLUDED							
(2) MW OF C ₆ +							

MATERIAL BALANCE

SENSITIVITY CASE I-B

-309-

STREAM N ^o		4.2	5.2	43.1	54	
		OXYGEN TO UNIT 116	PROCESS STEAM UNIT 116	PROCESS COND. FROM UNIT 116	LIGHT GAS TO SNG	
CO ₂	44.010 lbmol/hr				4.3	
H ₂ S	34.076 lbmol/hr					
COS	60.070 lbmol/hr					
C ₂ H ₄	28.054 lbmol/hr				8.7	
CO	28.010 lbmol/hr				220.6	
H ₂	2.016 lbmol/hr				9.2	
CH ₄	16.043 lbmol/hr				65.7	
C ₂ H ₆	30.070 lbmol/hr				57.4	
N ₂ +INERT	28.013 lbmol/hr	159				
O ₂	32.000 lbmol/hr	8219.38				
TOTAL DRY GAS (1)	lbmol/hr	8378.38			436.3	
H ₂ O	18.016 lbmol/hr		24 846	26275.08		
METHANOL	32.042 lbmol/hr					
TOTAL WET GAS (1)	lbmol/hr	8378.38			436.3	
TOTAL DRY GAS	lb/hr	267474			13317	
H ₂ O	lb/hr		447626	473372		
METHANOL	lb/hr					
C ₃ H ₆	42.08 lbmol/hr				1.4	
C ₃ H ₈	44.09 lbmol/hr				27.9	
iC ₄ H ₁₀	58.12 lbmol/hr				20.3	
1C ₄ H ₈	56.10 lbmol/hr				2.2	
nC ₄ H ₁₀	58.12 lbmol/hr				4.8	
iC ₅ H ₁₂	72.15 lbmol/hr				8.2	
1C ₅ H ₁₀	70.13 lbmol/hr				1.4	
nC ₅ H ₁₂	72.15 lbmol/hr				0.7	
C ₆ +	(2) lbmol/hr				3.5	
NAPHTHA		lb/hr				
TOTAL		lb/hr	267474	447626	473372	13317
(1) C ₃ + INCLUDED						
(2) MW OF C ₆ +						

MATERIAL BALANCE

SENSITIVITY CASE I-C

-310-

STREAM N°			38+35 - 33	54	39	44	48
			PURGE GAS TO SNG	LIGHT GAS TO SNG	FEED TO UNIT 112	UNIT 112 PRODUCT GAS	UNIT 113 OFFGAS
CO ₂	44.010	lbmol/hr	1483.00	44	1527.00	715.32	631.57
H ₂ S	34.076	lbmol/hr					
COS	60.070	lbmol/hr					
C ₂ H ₄	28.054	lbmol/hr	21.00	142	163.00		
CO	28.010	lbmol/hr	1268.80	89.5	1358.3	13.24	
H ₂	2.016	lbmol/hr	8288.54	253.2	8541.74	293.41	0.3
CH ₄	16.043	lbmol/hr	12450.63	179.2	12629.83	16060.26	16.3
C ₂ H ₆	30.070	lbmol/hr	275	18.4	293.4		
N ₂ +INERT	28.013	lbmol/hr	308.97		308.97	308.97	0.32
O ₂	32.000	lbmol/hr					
TOTAL DRY GAS (1)		lbmol/hr	24095.94	768.93	24864.87	17391.2	648.49
H ₂ O	18.016	lbmol/hr	3.00		3.0	15	45
METHANOL	32.042	lbmol/hr	233.00		233.00		
TOTAL WET GAS (1)		lbmol/hr	24331.94	768.93	25100.87	17406.2	693.49
TOTAL DRY GAS		lb/hr	334774	14218	348993	298754	28066
H ₂ O		lb/hr	54		54	270	811
METHANOL		lb/hr	7466		7466		
C ₃ H ₆	42.08	lbmol/hr		13.5	13.5		
C ₃ H ₈	44.09	lbmol/hr		29.13	29.13		
iC ₄ H ₁₀	58.12	lbmol/hr					
1C ₄ H ₈	56.10	lbmol/hr					
nC ₄ H ₁₀	58.12	lbmol/hr					
iC ₅ H ₁₂	72.15	lbmol/hr					
1C ₅ H ₁₀	70.13	lbmol/hr					
nC ₅ H ₁₂	72.15	lbmol/hr					
C ₆ +	(2)	lbmol/hr					
NAPHTHA		lb/hr					
TOTAL		lb/hr	342294	14218	356513	299024	28877
(1) C ₃ + INCLUDED							
(2) MW OF C ₆ +							

MATERIAL BALANCE

SENSITIVITY CASE I-C

-311-

STREAM N°			49	98	105		
			SNG PRODUCT	UNIT 150 HC PRODUCT	ALKYLATION FEED		
	CO ₂	44.010 lbmol/hr	83.75	44.0			
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr		142.7	0.7		
	CO	28.010 lbmol/hr	13.24	89.5			
	H ₂	2.016 lbmol/hr	293.11	253.2			
	CH ₄	16.043 lbmol/hr	16043.96	179.2			
	C ₂ H ₆	30.070 lbmol/hr		24.3	5.9		
	N ₂ +INERT	28.013 lbmol/hr	308.18				
	O ₂	32.000 lbmol/hr					
	TOTAL DRY GAS (1)	lbmol/hr	16742.24	4018.9	1521.4		
	H ₂ O	18.016 lbmol/hr	1.33				
	METHANOL	32.042 lbmol/hr					
	TOTAL WET GAS (1)	lbmol/hr	16743.57				
	TOTAL DRY GAS	lb/hr	270674				
	H ₂ O	lb/hr					
	METHANOL	lb/hr					
	C ₃ H ₆	42.08 lbmol/hr		362.8	349.2		
	C ₃ H ₈	44.09 lbmol/hr		240.5	237.1		
	iC ₄ H ₁₀	58.12 lbmol/hr		613.7	596.7		
	1C ₄ H ₈	56.10 lbmol/hr		253.6	230.5		
	nC ₄ H ₁₀	58.12 lbmol/hr		77.0	65.4		
	iC ₅ H ₁₂	72.15 lbmol/hr		446.5	26.6		
	1C ₅ H ₁₀	70.13 lbmol/hr		114.9	3.2		
	nC ₅ H ₁₂	72.15 lbmol/hr		22.6			
	C ₆ +	(2) lbmol/hr		1154.9			
	NAPHTHA	lb/hr					
	TOTAL	lb/hr	270698	258 303	79 270		
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						

MATERIAL BALANCE

SENSITIVITY CASE II-A

STREAM N°			30	37	38	44	49
			Feed to Unit 250	Unit 252 Feed	Unit 210 Product Gas	Unit 212 Product Gas	SNG Product
	CO ₂	44.010 lbmol/hr	2 900	978.22	2 132.90	890.87	85.00
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr	21	33.6	31.84		32.48
	CO	28.010 lbmol/hr	18 500	17.65	929.44	1.44	19.07
	H ₂	2.016 lbmol/hr	46 985	45.93	8 705.18	250	299.33
	CH ₄	16.043 lbmol/hr	12 460	1 070.36	14 007.74	17 306.45	18 326.20
	C ₂ H ₆	30.070 lbmol/hr	277	414.46	518.16		405.29
	N ₂ +INERTS	28.013 lbmol/hr	309	4.96	303.79	303.79	308.77
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr	81 452	3 995.69	26 634.05	18 752.55	19 528.74
	H ₂ O	18.016 lbmol/hr		113.98			
	METHANOL	32.042 lbmol/hr	1.31				1.70
	TOTAL WET FLOW (1)	lbmol/hr	81 453.31	4 109.67			19 530.44
	C ₃ H ₆	42.08 lbmol/hr		87.27			17.73
	C ₃ H ₈	44.09 lbmol/hr		273.13			34.47
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr		122.22			0.10
	iC ₄ H ₁₀	58.12 lbmol/hr		160.60			0.10
	nC ₄ H ₁₀	58.12 lbmol/hr		117.78			
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr		250.98			0.10
	nC ₅ H ₁₂	72.15 lbmol/hr		125.95			
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr		292.6	5.00		0.10
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	958 028	17 035	387 823		323 000
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

MATERIAL BALANCE

SENSITIVITY CASE II-A

-313-

STREAM N°			50	54			
			Gasoline Product	Unit 252 Off Gas			
	CO ₂	44.010 lbmol/hr		946.27			
	H ₂ S	34.076 lbmol/hr					
	COS	60.070 lbmol/hr					
	C ₂ H ₄	28.054 lbmol/hr		32.48			
	CO	28.010 lbmol/hr		17.63			
	H ₂	2.016 lbmol/hr		49.32			
	CH ₄	16.043 lbmol/hr		1 056.75			
	C ₂ H ₆	30.070 lbmol/hr		405.29			
	N ₂ +INERTS	28.013 lbmol/hr		4.98			
	O ₂	32.000 lbmol/hr					
	TOTAL DRY FLOW (1)	lbmol/hr		2 565.32			
	H ₂ O	18.016 lbmol/hr		113.38			
	METHANOL	32.042 lbmol/hr					
	TOTAL WET FLOW (1)	lbmol/hr	2 104.3	2 678.70			
	C ₃ H ₆	42.08 lbmol/hr		17.73			
	C ₃ H ₈	44.09 lbmol/hr		34.47			
	iC ₄ H ₈	56.10 lbmol/hr					
	iC ₄ H ₈	56.10 lbmol/hr		0.10			
	iC ₄ H ₁₀	58.12 lbmol/hr		0.10			
	nC ₄ H ₁₀	58.12 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₀	70.13 lbmol/hr					
	iC ₅ H ₁₂	72.15 lbmol/hr		0.10			
	nC ₅ H ₁₂	72.15 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₂	84.16 lbmol/hr					
	iC ₆ H ₁₄	86.17 lbmol/hr					
	nC ₆ H ₁₄	86.17 lbmol/hr					
	C ₆ +	(2) lbmol/hr		0.10			
	C ₇ +	(3) lbmol/hr					
	ALCOHOLS	46.61 lbmol/hr					
	TOTAL	lb/hr	186 556	76 800			
	(1) C ₃ + INCLUDED						
	(2) MW OF C ₆ +						
	(3) MW OF C ₇ +						

MATERIAL BALANCE

SENSITIVITY CASE II-A

STREAM N°		99	102	103	105	138
		Stabilized Gasoline	Unit 210 Feed	Unit 252 Feed	Acid Water from Unit 250	C ₃ LPG Product
	CO ₂ 44.010 lbmol/hr		3 111.12	15.88		
	H ₂ S 34.076 lbmol/hr					
	COS 60.070 lbmol/hr					
	C ₂ H ₄ 28.054 lbmol/hr		65.44	0.56		
	CO 28.010 lbmol/hr		947.09	0.91		
	H ₂ 2.016 lbmol/hr		8 751.11	5.88		
	CH ₄ 16.043 lbmol/hr		15 078.10	39.89		
	C ₂ H ₆ 30.070 lbmol/hr		932.62	11.38		
	N ₂ + INERTS 28.013 lbmol/hr		308.73	0.27		
	O ₂ 32.000 lbmol/hr					
	TOTAL DRY FLOW (1) lbmol/hr		30 629.69	912.05		
	H ₂ O 18.016 lbmol/hr		113.98		16 985.0	
	METHANOL 32.042 lbmol/hr					
	TOTAL WET FLOW (1) lbmol/hr	1 675.2	30 743.67			282.4
	C ₃ H ₆ 42.08 lbmol/hr		87.27	2.73		
	C ₃ H ₈ 44.09 lbmol/hr		273.16	9.84		
	iC ₄ H ₈ 56.10 lbmol/hr					
	iC ₄ H ₈ 56.10 lbmol/hr		122.22	10.80		
	iC ₄ H ₁₀ 58.12 lbmol/hr		160.60	12.40		
	nC ₄ H ₁₀ 58.12 lbmol/hr		117.78	12.22		
	iC ₅ H ₁₀ 70.13 lbmol/hr					
	iC ₅ H ₁₀ 70.13 lbmol/hr					
	iC ₅ H ₁₂ 72.15 lbmol/hr		250.98	57.52		
	nC ₅ H ₁₂ 72.15 lbmol/hr		125.95	34.45		
	iC ₆ H ₁₂ 84.16 lbmol/hr					
	iC ₆ H ₁₂ 84.16 lbmol/hr					
	iC ₆ H ₁₄ 86.17 lbmol/hr					
	nC ₆ H ₁₄ 86.17 lbmol/hr					
	C ₆ + (2) lbmol/hr		297.60	695.96		
	C ₇ + (3) lbmol/hr					
	ALCOHOLS 46.61 lbmol/hr					
	TOTAL lb/hr	157 045	557 858	94 100	306 070	12 419
	(1) C ₃ + INCLUDED					
	(2) MW OF C ₆ +					
	(3) MW OF C ₇ +					

APPENDIX D
SNG ONLY SENSITIVITY CASE

In the multiple products mode of economic evaluation, it is quite critical to employ the proper SNG value both for determining the effects of making motor gasoline (or methanol) and for calculating a correct gasoline (or methanol) price. Errors and wrong conclusions most likely will occur if the SNG value and plant complex used in the comparison are from the literature, where the coal and venture bases are different or unknown. Consequently, a SNG only sensitivity case has been added to this study. It was developed from the methanol Sensitivity Case I-A.

The results are summarized below:

PRODUCT YIELDS

SNG:	275 MMSCF/SD @ 983 Btu/SCF (HHV)
Hydrotreated Naphtha	1315 B/SD

INPUT

Net Coal to Gasifiers and Boiler:	24,070 T/SD
Water:	6,000 gpm

OVERALL PLANT EFFICIENCY (HHV)	68%
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PLANT COSTS

Investment	<u>MM \$</u>
Contractor Cost	1,083
Other Project Costs	221
Estimating Allowance	196
Royalties	17
Depreciable Capital	<u>1,517</u>
Expense Capital	28
Working Capital	44
Coal Cost	64.3 MM \$/yr
Operating Cost	120.1 MM \$/yr
Byproduct Credits	11.8 MM \$/yr

UNIT COSTS

Equity Financing	6.17 \$/MM Btu
Utility Financing	4.51 \$/MM Btu

The block flow diagram is similar to the one for Sensitivity Case I-A, except the methanol synthesis train is eliminated. In addition, because a balanced synthesis gas is not required for methanation as for methanol synthesis, the carbon dioxide can now be removed in the rectisol unit to the desired SNG level, thus eliminating the second CO₂ removal unit. To meet the methanation H/C ratio requirement, however, a higher percentage of the crude syn gas must be shifted then when synthesizing methanol.

Interestingly, even though the SNG and hydrotreated naphtha thermal yield is smaller than the SNG, methanol and hydrotreated naphtha thermal yield in Sensitivity Case I-A, the overall plant thermal efficiency is greater by about 2 percent. This improvement is the result of: (1) more efficient heat recovery - greater 100 and 550 psig steam production - in the shift and methanation units, (2) elimination of the steam requirements for methanol/water fractionation, and (3) lower compressor power requirements.

The SNG unit cost is a stand-off with the thermal cost for SNG/methanol coproduction, 6.17 vs. 6.19 \$/MM Btu @ 12% DCF. The economics of the SNG case are helped by lower net coal usage (higher efficiency), lower plant investment (smaller boiler), lower working capital (no product inventory), and lower operating cost.

The effect of the product mix on the plant thermal balance and economics is briefly summarized in the attached table.

EFFECT OF PRODUCT MIX ON PLANT
THERMAL BALANCE AND ECONOMICS

<u>Case</u>	<u>SNG Only</u>	<u>I-A SNG + MeOH</u>	<u>Base Case I SNG + Gasoline</u>	<u>I-B Gasoline Only</u>
Thermal Balance				
Net Coal, MM Btu/hr	17,067	18,280	18,511	19,458
Product, MM Btu/hr	11,545	11,970	11,388	9,050
Efficiency	68	66	62	47
Total Investment, MM \$	1,589	1,637	1,790	2,062
Operating Cost, MM \$/yr	120	122	132	149
Unit Cost				
Equity, \$/MM Btu	6.17	6.15	6.99	9.98
Utility, \$/MM Btu	4.51	4.47	5.08	7.24