APPENDIX B
TRANSPORT LOGISTIC STUDY

APPENDIX B

LOGISTICS STUDY

The objectives of this study were to determine the optimum means of distributing products produced at three liquefaction sites to designated marketing areas and to determine the relative cost between the marketing areas for distribution charges for products from the five highest commercial potential processes. The following were determined:

- Whether pipeline distribution is the best means of delivery
- What pipelines now exist which can be used for this purpose
- What pipelines and supporting facilities would have to be constructed to achieve pipeline distribution
- The total costs of distribution according to recommended distribution plans, including amortization of construction costs of any new facilities required

Three liquefaction plant locations were selected for use in this study:

1. Gillette, Wyoming
2. New Athens, Illinois
3. Moundsville, West Virginia

Plant products generated in Gillette, Wyoming would be delivered to markets primarily in the San Francisco area. Plant production generated in the New Athens, Illinois plant would be delivered primarily to Chicago, with

Houston as a second market center. Plant products from the Moundsville, West Virginia plant would be delivered to Delaware.

Four coal liquefaction processes have been selected as high potential processes. These are:

1. SRC II Based
2. H-Coal Based
3. CSF Based
4. Fischer-Tropsch Based

B.1 PLAN OF APPROACH

Pipeline, railroad and barge transportation will be considered for moving products from each plant site to its marketing center or secondary marketing center. Where a choice of mode of transportation exists, the lowest cost method will be assumed.

B.1.1 PIPELINES

In order to assemble and evaluate the data and information required in the study, the following approach was taken.

By library research and by contact with pipeline companies, the tariff rates and availability of pipelines now existing between plant sites and marketing areas was determined. The line size, capacity, auxiliary storage facilities and direction of flow were also obtained as important related information. In some cases, even though the line may be near capacity at present, the rates were recorded for two reasons:

1. The petroleum product distribution patterns may change significantly over the years while the plants are being designed and constructed.

2. The rates provide a comparative distribution cost which can be used to evaluate the relative efficiency of pipelines even if they are not available for use at present.

We have considered Houston, Texas as a secondary marketing center for products produced at New Athens, Illinois, but at present all product pipelines flow in the opposite direction from Houston to the rest of the United States. Again, this may not be the situation by the time the liquefaction plants are operating, and it has been assumed that the tariff, say, from Houston to St. Louis will be about the same if materials were moving in the other direction to Houston; therefore this rate was included in the study for comparative purposes.

As interstate carriers, the pipelines and the tariffs that they can be charged are controlled by the Federal Government and the rates include cost of amortizing pipeline construction costs over a period of years which may relate to the estimated length of time the supply will last. The liquefaction plants will be a new source of supply to be developed as existing petroleum sources diminish. Therefore, liquefaction plant products may move through pipelines which have been fully amortized, and such movement will be at greatly reduced rates. Since pipelines are the least expensive means of moving most products, primary emphasis has been placed on this transportation mode in this study. When no pipeline could be found to move the products generated to the designated market areas, the cost of constructing a pipeline was estimated and amortized to provide a basis for daily operating costs. In some cases, we considered that half of this amortized cost could be shared with other companies.

B.1.2 RAILROADS

In the case of each primary market area for the three liquefaction plant sites, the tariff structure for the products was determined. Rates for the largest tank cars were used, but high volume shippers can get special tariffs approved and obtain some reductions from the quoted rates. Exactly how much reduction would be possible cannot be determined at this time. It is also possible that reduced rates can be obtained by using unit trains made up of tank cars, but no quotations of such rates are available. In every case it was apparent that pipeline rates were substantially below railroad rates.

B.1.3 BARGE TRANSPORTATION

In the case of movement of Gillette, Wy. products to San Francisco, no barge rates were obtained because of lack of proximity to suitable waterways. Also, at the Moundsville, W. Va. site there is no direct access by water to Dover, Delaware. The Appalachian Mountains are between Pittsburgh and Philadelphia, obstructing barge movement. The two possible routes to follow are:

1. Down river to New Orlenans, transfer to seagoing ships to move north and up the Delaware Bay.
2. Up river to Chicago, through the Great Lakes, transfer to seagoing ship to move south and up the Delaware Bay.

Both of these routes are considered to be too circuitous and expensive to be competitive considering the much shorter direct distance between Moundsville and Dover.

Barge rates were obtained for the movement of products from St. Louis to Chicago, and they were found to be less than pipeline rates for LPG. Also, barge transportation appears to be the best means of transportation between St. Louis and Houston for LPG and residual fuel oil.

B.1.4 SUPPLEMENTAL AND NEW CONSTRUCTION REQUIRED

In each case when a pipeline company was contacted regarding the movement of LPG (propane-butane), a rate was quoted with the stipulation that storage tanks be provided at the shippers' expense. Thus, the cost of such tanks has been included. Also, backup tanks have been provided at the plant site so that minor breakdown problems in transportation will not shut down operations.

B.2 COST DERIVATION AND COMPARISONS

B.2.1 NEW ATHENS, ILLINOIS TO CHICAGO AND HOUSTON

B.2.1.1 New Athens, Illinois to Chicago, Illinois

Pipelines were selected to transport SNG, gasoline and No. 2, No. 4 and No. 6 fuel oil from New Athens to Chicago. Barge delivery was least expensive for LPG and oxygenates. Relative transportation costs for the total product slate of each of the four selected processes are shown in Table B-1. These transportation cost ratios are based on a value of 1.00 for the cost of transportating SRC II products from New Athens to Chicago.

B.2.1.2 New Athens, Illinois to Houston, Texas

Pipelines were selected to transport SNG, gasoline, No. 2 fuel oil and No. 4 fuel oil from New Athens to the secondary marketing center of Houston. Barge delivery was chosen for LPG, No. 6 fuel oil and oxygenates. Transportation cost ratios are shown in Table B.1.

B.2.2 MOUNDSVILLE, WEST VIRGINIA TO DOVER, DELAWARE

Pipelines were assumed for all product transportation from Moundsville to Dover. Railroads were more expensive for all products and barge transportation would be too circuitous and expensive. Transportation cost ratios are shown in Table B-1.

B.2.3. GILLETTE, WYOMING TO SAN FRANCISCO, CALIFORNIA

Pipelines were less expensive than railroad transportation for all products and no barge transportation is available. Transportation cost ratios are shown in Table B-1.

B.3 ALTERNATIVE PRODUCT DISPOSAL

Houston is now a productive refining center with a vast product distribution system radiating to the east, north and west. With the sources of oil in Texas diminishing, the volume of flow of this distribution system may also diminish. A reverse flow from inland coal refineries is conceivable. San Francisco, Dover and Houston are all ocean front locations where imported oil and oil products have easy access. On the other hand, the Rocky Mountain States, the Appalachian States and Illinois, are all in the interior of the U.S. where foreign oil imports are subject to high transportation costs. The most logical distribution scenario would appear to be to obtain maximum distribution of liquefaction products at the market areas in the vicinity of plant location, and move only the excess to distant metropolitan centers where these products will be less profitable due to high transportation costs. There are substantial markets in Salt Lake City, St. Louis and Pittsburgh, which can absorb most of the planned liquefaction output with minimum transportation costs.

Another product disposal plan which has several advantages is that of selling products at the plant to oil and gas companies for distribution and retail sale. Thus, the financing and construction of pipelines, tanks, etc., becomes the concern of the distributors. And, of course, the strong possibility of product trades exists which would greatly reduce some of the product transportation charges.

B.4 CONCLUSIONS

With reference to the objectives of this study, it is concluded that:

- Pipeline distribution is the most economical means of distributing liquefaction plant products, except in those cases where a river system is adjacent to both the plant and the market area. Where barge movement is possible, it is less expensive in some cases than pipeline transportation. These factors are not absolute due to equipment amortization as it affects the allowable rate charged.

- An extensive pipeline system now exists in the U.S. for both petroleum products and natural gas distribution. At present many of these lines have inadequate surplus capacity available to move the liquefaction products to be generated. From Houston, Texas the flow is in the opposite direction. However, these factors may be mitigated by the time liquefaction plants are built. In several cases pipeline companies contacted expressed an interest in purchase of the products at the plant site for distribution and resale to customers at their expense.

B.5 RECOMMENDATIONS

It is recommended that:

1. Further studies be conducted of the most probable spectrum of petroleum product needs in the several U.S. marketing areas being considered at the time the liquefaction plants will be in operation.

2. Further refined cost estimates be developed for the new construction required to accomplish distribution, with emphasis on pipelines and supporting equipment.

3. Consideration be given to distribution modes with the highest profit margin return for the products generated, considering that imported petroleum products will have easiest access to costal cities.

Table B-1 - Transportation Cost Ratios

Manufacturing Location	Marketing Center	Process			
		SRC II Based	H-Coal Based	CSF Based	F-T Based
New Athens, Ill.	Chicago, Ill.	1.00	1.14	1.05	2.07
New Athens, Ill.	Houston, Tex.	2.11	2.24	2.20	2.50
Moundsville, W. Va.	Dover, Del.	2.17	2.16	2.19	1.67
Gillette, Wy.	San Francisco, Calif.	4.20	4.38	4.28	4.87

APPENDIX C
M-GASOLINE BASED PROCESS

APPENDIX C

M-GASOLINE BASED PROCESS

Preliminary technical and relative economic factors are summarized here for the M-Gasoline based process. It obtained a high rating in the preliminary assessment described in APPENDIX A and drew special interest by DOE. However, because the evaluation basis used for the then existing Mobil-M assessment was established for another study[22,24] and differed from that used for the comparison of SRC II, H-Coal, CSF and Fischer-Tropsch, the Mobil-M results are summarized here separately for reference. Future comparisons should include Mobil-M as a candidate.

C.1 DEVELOPMENT STATUS

The process consists of two major steps which are:

1. The production of methanol from coal
2. The production of gasoline from methanol

The production of methanol from coal is practiced commercially in a number of countries of the world and has been practiced on an industrial scale in the U.S. The process and its economics are dominated by the process performance in production of synthesis gas from coal.

The conversion of methanol to high octane gasoline is under development by the Mobil Research and Development Corporation. Work has included laboratory scale and a four-barrel-per-day Process Development Unit (PDU) operated by Mobil at Paulsboro, New Jersey. DOE, and its predecessor ERDA, have supplied funds to support this work.

C.2 DATA SOURCES AND FIXED CAPITAL INVESTMENT BASES

The data for the coal to methanol process and its yields were taken from a report prepared for ERDA by the E. I. Dupont de Nemours Company.[22] Adjustments were made to the design and economics to correct for the fact that the coal composition used in the reference report differed from that used in this report.

The data for the methanol to M-Gasoline process and its yields were taken from a report prepared for ERDA by the Mobil Research and Development Corporation.[23]

The bases for the fixed capital investments for the plant were taken from the two indicated references.[22,23]

C.3 PROCESS DESIGN BASIS

The process concept and economics for the M-Gasoline process were developed under guidelines which differed from those prescribed for the SRC II, H-Coal, donor solvent and Fischer-Tropsch processes as covered in Section 5 of this report. Therefore, the results are not directly comparable. A key difference is the selection of gasifier; the first four process concepts included second-generation pressurized entrained gasifiers feeding either solid coal or coal-in-oil slurry. The M-Gasoline process, by contrast, used a coal-in-water slurry feed with its attendant decrease in thermal efficiency. There are a number of other differences in design philosophies and concepts between the M-Gasoline concept which follows and the other four processes.

C.4 PROCESS DESCRIPTION

The M-Gasoline process is shown schematically in block flow diagram, R-05-FS-1.

For ease of presentation, the process has been divided into three sections:

1. Conversion of coal to a clean, low sulfur synthesis gas
2. Conversion of syngas to crude methanol
3. Conversion of crude methanol to gasoline and LPG products

In the first section, prepared coal is slurried with water and pumped to several entrained Texaco type gasifiers which operate at about 2,300°F and 800 psig. The synthesis gas is quenched and cooled while generating high pressure steam. Dust removal is accomplished in a three-stage dust removal section including a moisture saturation process, a Venturi scrubber, and a gas/solids removal wash tray. Sour synthesis gas, after dust removal and final quenching, is shifted to the required H_2/CO ratio for methanol synthesis. The H_2S is removed in the selective H_2S removal unit which is the first stage of the Selexol process; the second stage removes the CO_2 and COS from the shifted synthesis gas. The CO_2 rich gas is used to strip entrained gases from slag produced in the gasifiers.

The syngas, now at about 680 psig and 85°F from the CO_2 removal unit, enters the second section of the plant (conversion to methanol). It is first passed through desulfurizer drums which contain activated carbon granules that act as a sulfur guard for the methanol synthesis catalyst. The sulfur-free syngas is compressed to about 1,500 psig, preheated, and passed through several catalyst beds to convert the syngas to methanol. Recycled quench gas is used to dissipate the heat released in the reaction.

To separate the unreacted gases from the product methanol, the reactor outlet gas is cooled in a heat exchanger train. The resulting two-phase mixture produced by cooling is separated and the unreacted gases are compressed. Dimethylether and higher alcohols are collected and sent, with methanol, to the methanol-to-gasoline conversion unit (M-Gasoline plant). Purge gas from methanol synthesis is used for steam generation.

In the third section of the plant, methanol is converted to gasoline, LPG, and a fuel gas. Crude methanol is preheated to 600°F at 300 psig and the methanol, as a vapor, enters the first of a two-stage reaction system. The first reaction catalytically converts the methanol to an equilibrium mixture of dimethylether, methanol and water.

The reaction is exothermic; hence, the effluent leaves the reactor at about 770°F. The gas from this reactor is cooled to 650°F by quenching with a cool recycle gas before entering one of several reactors in parallel required for the second reaction. In this second reaction, a shape-selective zeolite catalyst is used to complete the conversion of methanol to hydrocarbons and water. The overall reaction can be represented as follows:

$$nCH_3OH \qquad (CH_2)n + nH_2O$$

Essentially no hydrocarbons having more than eleven carbon atoms are produced.

Condensed water is treated and used for cooling tower make-up and coal slurrying. Hydrocarbons produced from the reaction are separated in the gas plant to produce SNG, a C_3-C_5 liquids stream as depentanizer overhead, and C_6-plus liquids.

The depentanizer overhead, containing isobutane, is depropanized and sent to an HF alkylation unit to alkylate the butenes and pentenes. The alkylate is blended with the sythesized C_6-plus liquids and butanes from the M-Gasoline reaction to produce M-Gasoline product.

A coal-fired steam plant is provided to produce the plant steam requirements to supplement that produced by waste heat boilers in the gasifier and shift areas. In addition to firing coal, the steam plant boilers also fire spent char from the gasifier, stripped gases from slag removal, tail gas from the sulfur plant, purge gas from methanol synthesis, fuel gas from the gas plant, and regeneration gases from the M-Gasoline plant. The flue gas clean-up unit is provided to collect the fly ash and unburned carbon from the boiler flue gas (utilizing wet high energy collection) followed by SO_2 absorption from the gas and then SO_2 recovery in concentrated form. The SO_2 absorption is accomplished using a regenerative Wellman-Lord process; most of the SO_2 recovered is sent to the sulfur plant where it combines with the H_2S from the selective H_2S removal unit to produce elemental sulfur. The rest of the SO_2 from the boiler flue gas is recovered as Na_2SO_4 in the flue gas clean-up unit.

The plant is a self-contained facility producing all of its own power, steam, and fuel gas.

C.5 ASSESSMENT OF PROCESSING OTHER COALS

In the M-Gasoline process, as in the Fischer-Tropsch process, all the coal fed to the process is gasified to produce a synthesis gas which is converted to the final products. In this report the M-Gasoline process is based on the use of the Texaco gasifier and the Fischer-Tropsch process uses the BCR gasifier concept. Although different types of gasifiers are used, the trends in yield pattern resulting from use of different coals should be similar. The information presented in Table 5-3 can therefore be used to indicate the impact on the M-Gasoline process of using alternate coals even though the relative changes when using the Texaco gasifier may be different.

As previously shown for the Fischer-Tropsch process, the M-Gasoline process would require less coal feed and less oxygen to produce the same quantity of products from Eastern Coal Province coal. This should translate into a decrease in the capital cost since both the coal preparation area and the oxygen plant would be smaller. There should be little change in plant thermal efficiency since the decreased coal feed rate is offset by the higher HHV of the Eastern coal.

Again, as was shown for the Fischer-Tropsch process, the M-Gasoline process would require more coal feed and more oxygen to produce the same quantity of products from Rocky Mountain Coal Province coal. Therefore, the net change would be an increase in capital cost and a decrease in thermal efficiency. However, the cost of Rocky Mountain coal is expected to be significantly lower than the Eastern and Interior coals and would tend to compensate for any increased capital cost.

C.6 ENVIRONMENTAL CONSIDERATIONS

The major environmental aspects relating to the M-Gasoline process were considered in detail and are presented below.

C.6.1 AIR EMISSIONS

The generation and control of air emissions for the M-Gasoline plant are outlined in Figure C-1, which also shows the nature and amount of the streams vented to the air. All effluents are conveyed to the steam and power generation plant and are exhausted to the air with the flue gas.

Coal is fed to the process gasifier as a water slurry, and all grinding operations occur in a wet mode; therefore, no particulates are emitted during coal preparation.

The gas stream produced by the gasifier is freed of particulates by an abatement system consisting of a condensation scrubber followed by a Venturi scrubber and a wet tray scrubber; the separated char particles are used as fuel for the steam and power generating plant. After shift conversion, the gas stream is conveyed to an acid gas removal unit where the acid gases present (mainly hydrogen sulfide, hydrogen cyanide, carbon oxysulfide, and carbon dioxide) are separated from the other main syngas components (hydrogen, carbon monoxide, and methane). On regeneration of the solvent, carbon dioxide with traces of carbon monoxide and hydrogen sulfide is separated from the other acid gases and added to the combustion air fed to the steam and power generating plant. Hydrogen sulfide and other contaminants are conveyed to the sulfur plant.

The last traces of sulfur present in the cleaned synthesis gas are removed by an activated carbon guard bed, to avoid deactivating the methanol and gasoline synthesis catalysts. The liquid fuels produced are essentially nitrogen and sulfur free.

The sulfur plant converts approximately 95 percent of the hydrogen sulfide to elemental sulfur; other contaminants are thermally oxidized to inoffensive compounds. The tail gas from the sulfur plant joins the combustion air fed to the steam and power generating plant.

Steam and power are generated by combustion of crushed coal, char separated from the process gasifier gas stream, and fuel gas obtained as by-products from hydrocarbon synthesis. The solid fuels contain substantial amounts of sulfur, requiring flue gas scrubbing for removal of sulfur dioxide. This is achieved in a Wellman-Lord flue gas desulfurization unit, after removal of particulates by a high energy wet scrubber. Desulfurization occurs by contacting the flue gas with a solutiion of sodium sulfite, which absorbs SO_2, becoming sodium bisulfite. On heating, sodium sulfite is recovered and recycled, while the SO_2 stream generated is conveyed to the sulfur plant; there sulfur dioxide reacts with hydrogen sulfide, producing elemental sulfur. During flue gas scrubbing, some sodium sulfite is oxidized to sodium sulfate, which is not reactive and is removed as a waste product. Part of the SO_2 is emitted to the air, together with carbon dioxide, particulates, nitrogen oxides, and small amounts of sulfur trioxide.

The amounts and concentrations of the effluents emitted to the air are also detailed in Table C-1, while the sulfur balance for the plant is presented in Table C-2. Compliance with applicable standards is discussed in the following section.

C.6.2 COMPLIANCE WITH APPLICABLE SOURCE AIR EMISSION STANDARDS

The projected M-Gasoline plant particulate emission rate is 175 lb/hr, which exceeds the emission standards allowed by the states of Illinois and Wyoming of 85 and 59 lb/hr, respectively. The sulfur dioxide emission standard for the state of Wyoming is also exceeded, emission being projected to be 0.26 lb/MM Btu fired against an allowable 0.20 lb/MM Btu fired. Both these excessive emissions are produced by the steam and power generating unit. This unit is similar to a coal burning power plant; many of these plants are presently operating in compliance with applicable emission standards. It is believed that upgrading of the air pollution control technology would provide the improvements required to meet the standards.

C.6.3 WATER TREATMENT AND SUPPLY

The generation and treatment of aqueous contaminants for the M-Gasoline plant is outlined in Figure C-2, which shows the treatment and disposition of the aqueous effluent streams; approximate flow rates are reported. Raw river water is purified by flocculation and settling and used for cooling water makeup and, after further sand filtration and deionization, for boiler feedwater makeup. The boiler blowdown is used for cooling water makeup.

The most highly contaminated stream is the sour water generated by several process units, such as the gasifiers and product distillation units. This stream contains hydrogen sulfide, ammonium sulfide, phenols, xylenols, thiocyanates, cyanides and solids (ash and char). Treatment consists of steam stripping to remove hydrogen sulfide and ammonia; the aqueous stream is then used for slag and ash quenching and disposal. Other more lightly contaminated streams, such as the oily water (after oil-water separation), deionizer wastewater, and water treatment sludges are also used for ash disposal. Cooling water blowdown is used for wet scrubbing of the flue gas from the steam and power generating unit, then is conveyed to the ash disposal unit. The overflow from the latter unit is used to slurry the coal fed to the gasifier.

The water treatment and supply system described is a "zero discharge" system, i.e., no water is returned to the river. However, most effluent streams are brought into contact with slag and ash; these materials, therefore, will become enriched in contaminants and require careful disposal to avoid leaching by percolation or runoff.

C.6.4 WATER CONSUMPTION

Water consumption is at the rate of 42,000 gpm. This is approximately twice that for the CSF process, which has the highest water requirement of the four processes assessed in the body of this report. Since this loss is primarily in the form of cooling water makeup, the need for improvement in heat recoveries in the M-Gasoline process is indicated. In a semi-arid location greater use of air coolers as well as improved heat recovery should be applied to the plant design.

C.6.5 GENERAL

Solid waste problems are expected to be similar to that of any average equivalent sized coal conversion plant.

Biohazards are expected to be minimal compared to plants having dissolver operations. No coal-derived carcinogenic hazards, either in the in-plant streams or final products are expected.

C.7 ECONOMICS

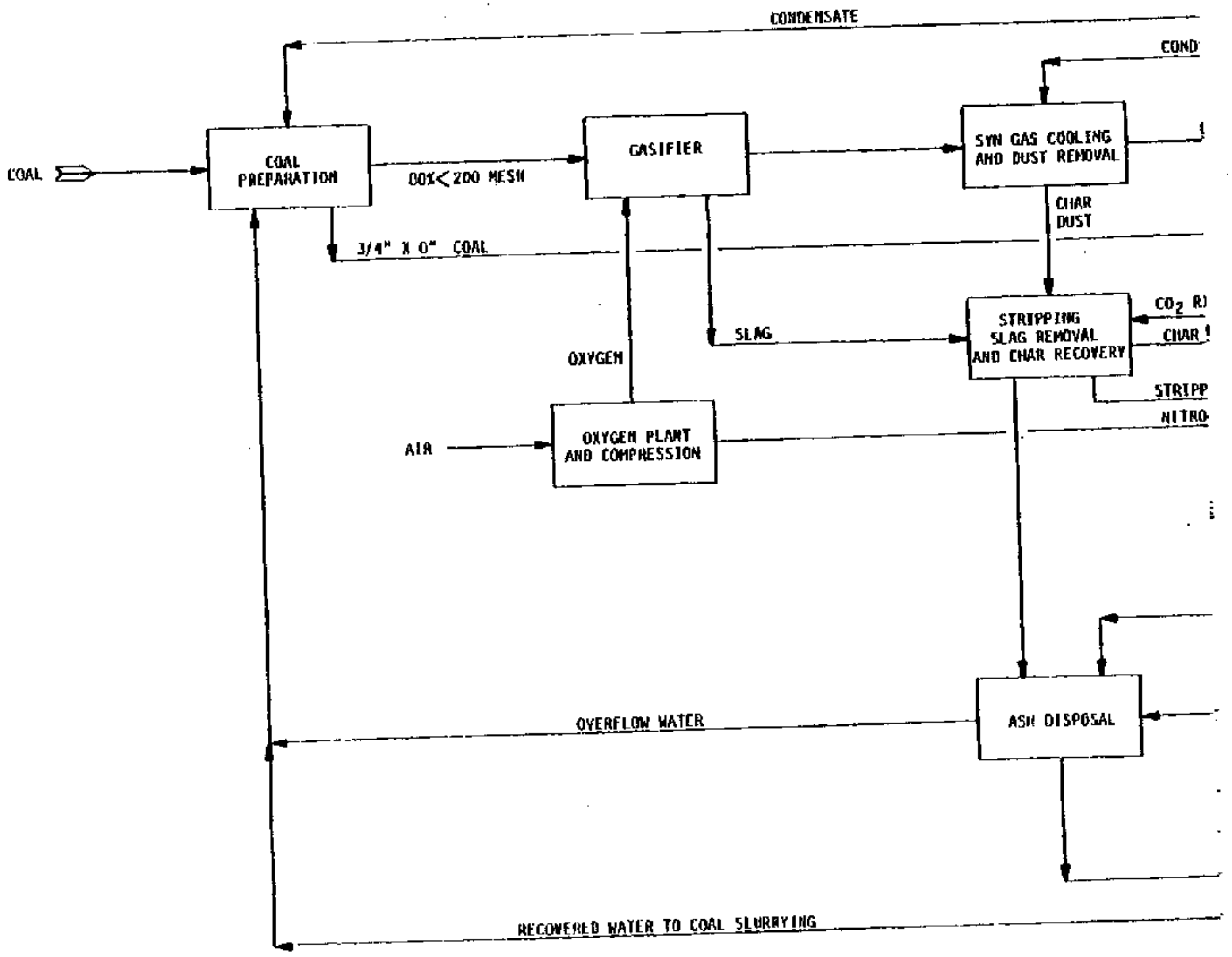
The economic procedures presented in Section 7 involved in the detailed analyses of the four high potential processes were also applied to the M-Gasoline process, also sized for a capacity of 600 billion Btu per day of products. The possible market values of the products are similar to the values contained in Table 7-1. A summary of pertinent analytical/economic information for base case location (Eastern Region of the Interior Coal Province) and also the two alternate locations (Appalachian Region of the Eastern Coal Province and the Powder River Region of the Rocky Mountain Province) are presented in Tables C-3 through C-8. All economics are presented in first quarter 1978 dollars and the same ratio system adopted in Section 7 is used; see Tables 7-2 through 7-6, inclusive.

Table C-3 summarizes the feed coal and product quantities for the base case (Eastern Region of the Interior Coal Province). The energy quantities given in C-3 indicate an overall thermall efficiency of 46 percent for the process on which the study is based.[22,23] In a study by The Ralph M. Parsons Company,[24] which consisted of a conceptual design, an overall thermal efficiency of 59 percent was indicated to be feasible. Higher product yields were indicated to be possible.

The fixed and total capital investment ratios are summarized in Tables C-4 and C-5 for the 107,000 BPSD gasoline plant. The cost of future plants might be reduced by the use of fewer and more advanced gasifiers and modifications to the methanol synthesis process.[24]

Table C-6 summarizes the operating cost ratios and Table C-7 summarizes the required product selling price ratios and the contributions of economic categories such as capital burden, cost of coal, operations cost and by-product credit. Projected total annual revenues ratios as well as required annual revenues and benefit/cost (B/C) ratio are shown in Table C-8.

Tables C-6 and C-7 indicate the methanol process configuration considered here is a potentially high cost synthetic gasoline producer. Techniques for improving the efficiencies and economics should be considered.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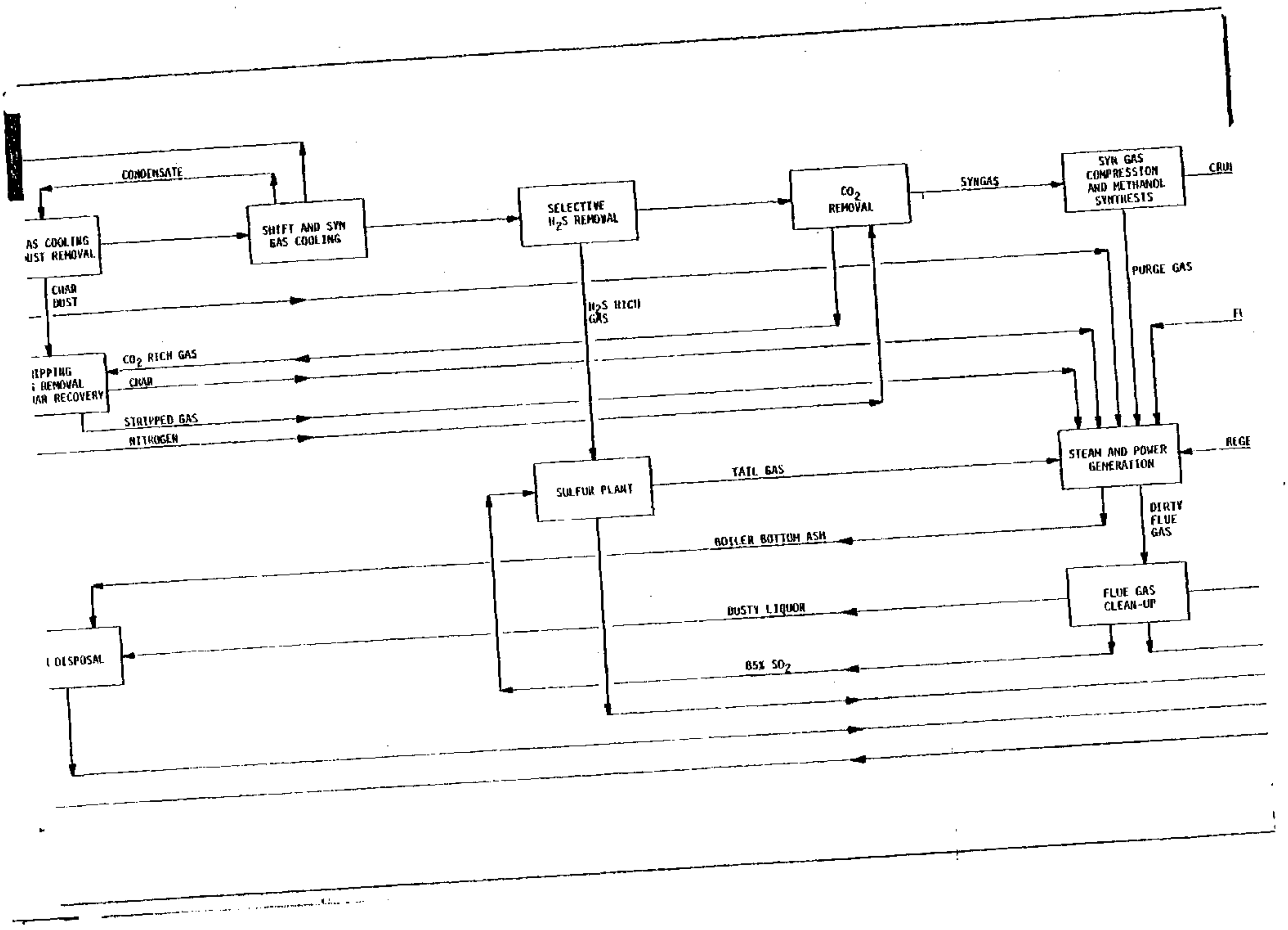
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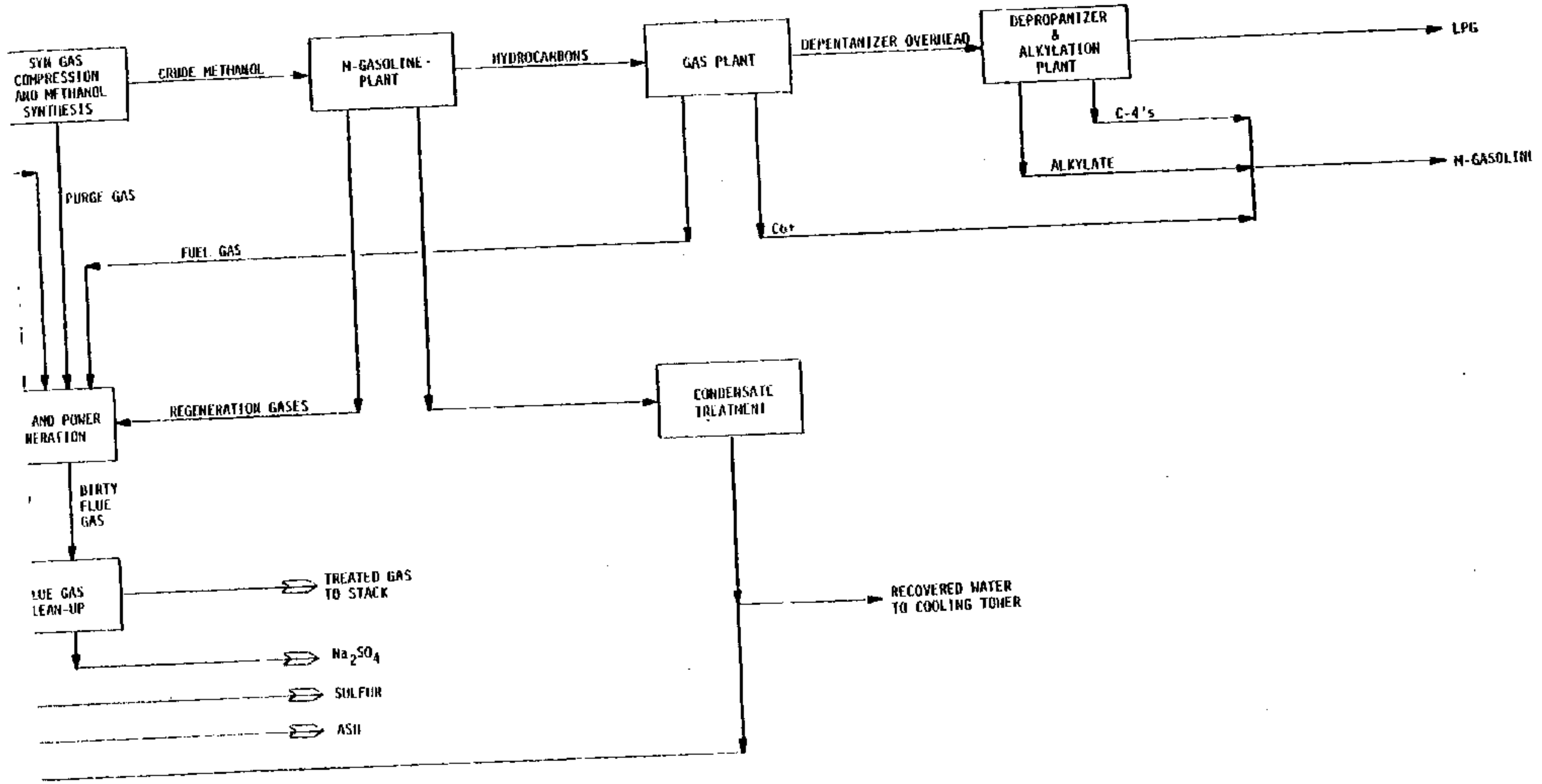
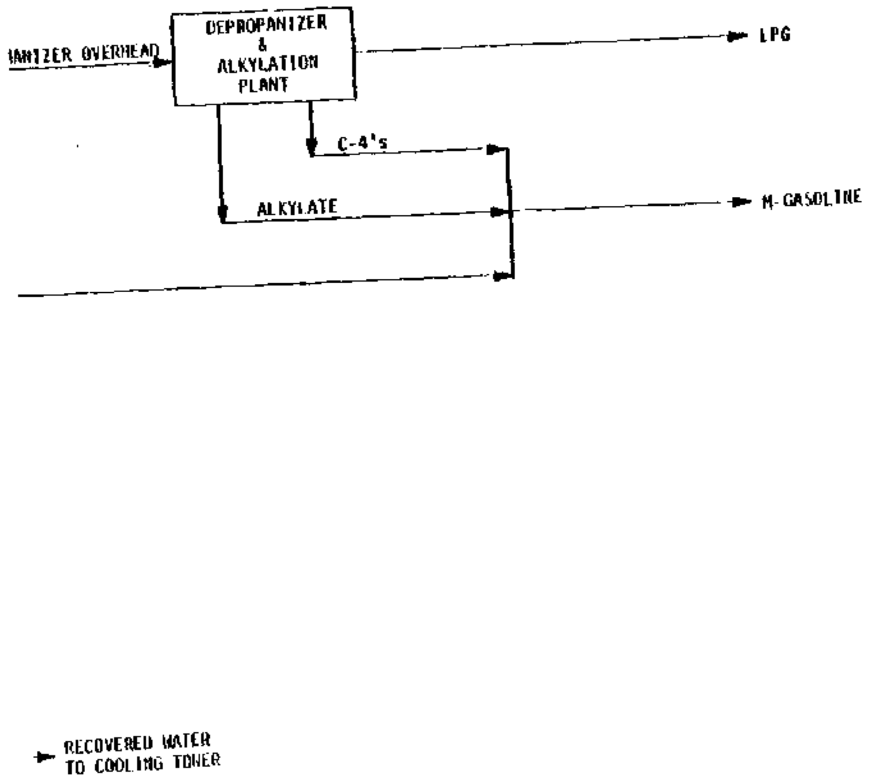
STRIPPED GAS
NITROGEN
SULFUR PLANT
TAIL GAS
STEAM AND POWER GENERATION
REGE
DIRTY FLUE GAS
BOILER BOTTOM ASH
FLUE GAS CLEAN-UP
DUSTY LIQUOR
85% SO_2
DISPOSAL


SYN GAS COMPRESSION AND METHANOL SYNTHESIS
CRUDE METHANOL
M-GASOLINE PLANT
HYDROCARBONS
GAS PLANT
DEPENTANIZER OVERHEAD
DEPROPANIZER & ALKYLATION PLANT
LPG
C-4's
ALKYLATE
M-GASOLIN
C6+
PURGE GAS
FUEL GAS
AND POWER
NERATION
REGENERATION GASES
CONDENSATE TREATMENT
DIRTY FLUE GAS
LUE GAS
LEAN-UP
TREATED GAS TO STACK
RECOVERED WATER TO COOLING TOWER
Na_2SO_4
SULFUR
ASH



	TONS/DAY	BPSD	MMBTU/D (HHV)
FEED COAL	51,630	-	1,295
PROPYLENE/PROPANE	1,140	12,790	49
GASOLINE	13,790	106,570	551
SULFUR	1,740	-	-
Na_2SO_4	25	-	-
TOTAL	16,695	119,360	600

THERMAL EFFICIENCY: $\frac{600}{1295}$ X 100 = 46%

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COAL LIQUEFACTION PROCESS SURVEY
BLOCK FLOW DIAGRAM
M-GASOLINE PROCESS

THE RALPH M PARSONS COMPANY
PASADENA, CALIFORNIA

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DWG NO R-DS-FS-1
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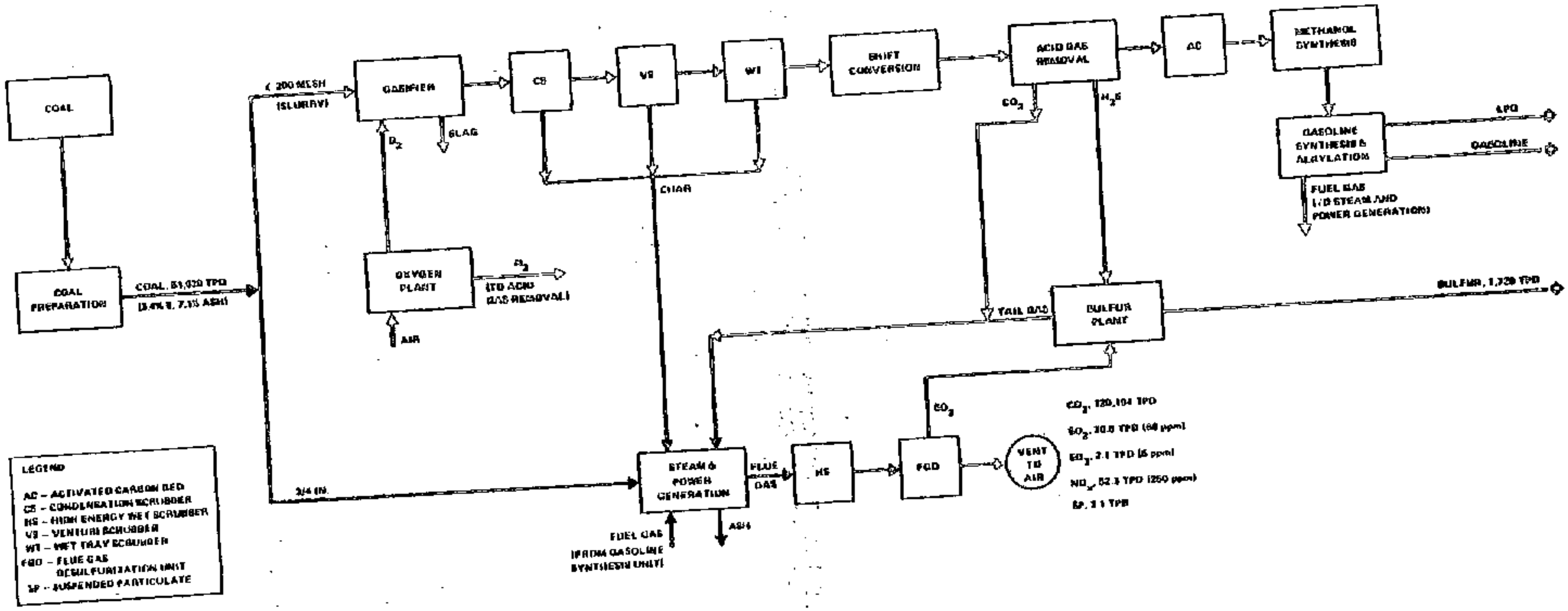


Figure C-1 - Block Flow Diagram,
Air Pollution Abatement, M-Gasoline Process

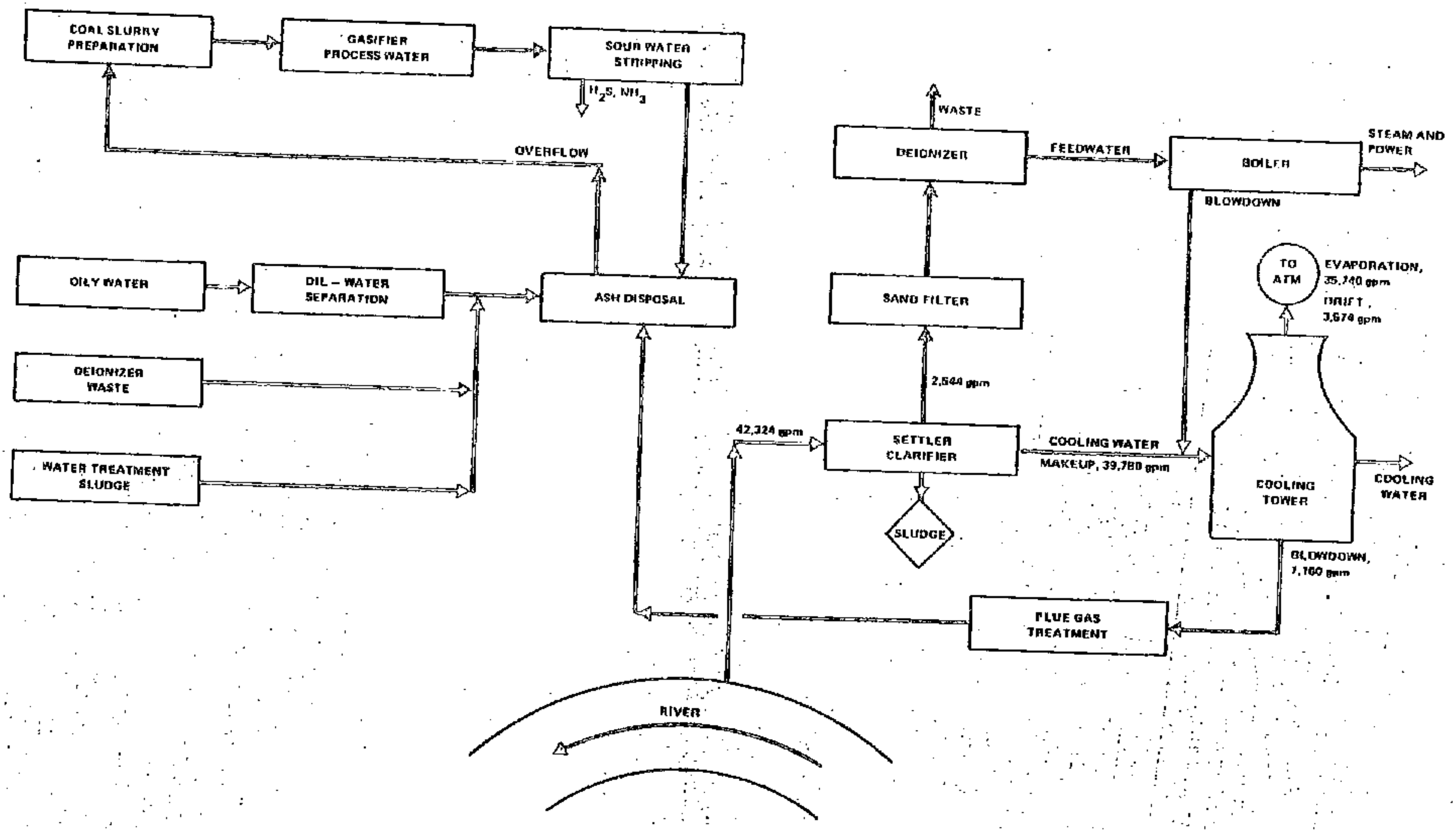


Figure C-2 - Block Flow Diagram, Water Treatment and Supply, M-Gasoline Plant

Table C-1 - Combined Effluents Emitted to the Air, M-Gasoline Based Plant

Outlet	Gaseous Effluent	Amount (TPD)	Concentration (ppm)
Plant Stack	Carbon Dioxide	120,193.0	-
	Sulfur Dioxide	20.8	60.0
	Sulfur Trioxide	2.1	4.8
	Nitrogen Oxides (as NO_2)	63.3	250.0
	Particulate Matter	2.1	-

Table C-2 - Sulfur Balance, M-Gasoline Based Process

Item	TPD
Total Input from the Typical Feed Coal	1755.0
Outputs:	
Recovered as Elemental Sulfur	1738.6
As Sulfur Oxides Emissions from Combustion of Solid and Gaseous Fuels for Steam and Power Generation	11.2
As Sodium Sulfate from Flue Gas Scrubber	5.2
Total Outputs	1755.0

Table C-3 - Coal Feed and Product Quantities

Item	Eastern Region - Base Case		
	TPD	BPSD (10^6 SCFD)	10^9 Btu/D
Feed Coal	51,650		1,296
Total Products			
Propylene/Propane	1,140	12,789	49
Gasoline	13,790	106,572	551
Sulfur	1,739		14
Na_2SO_4	33		
Total Products	16,702	119,361	614
Fuel Products			
MM Btu/ton of Coal		11.62	
MM Btu/day		600	
Liquid Products			
bbl/ton of Coal		2.31	
bbl/day		119,360	

Table C-4 - Fixed Capital Investment Ratio

Item	Relative Contribution to FCI (Eastern Region - Base Case)
Coal Preparation, Storage, Grinding and Drying	0.10
Process Operations	1.52
Steam and Power Generation	0.25
Offsites and General Facilities	0.27
Total Constructed Cost	2.14
Engineering, Home Office, Fee and Sales Tax at 12%	0.26
Fixed Capital Investment Ratio	2.40

Table C-5 - Total Capital Investment Ratio

Item	Relative Contribution to Total Investment		
	Eastern Region	Appalachian Region	Western Region
Fixed Capital Investment	1.89	1.87	1.84
Initial Catalysts	0.02	0.02	0.02
Startup Costs	0.04	0.04	0.04
Interest During Construction	0.24	0.23	0.23
Commitment Fee	0.04	0.04	0.04
Total Depreciable Investment	2.23	2.20	2.17
Working Capital	0.17	0.18	0.13
Total Capital Investment Ratio	2.40	2.38	2.30

Table C-6 - Operating Costs Ratio

Item	Relative Contribution to Operating Cost		
	Eastern Region	Appalachian Region	Western Region
Coal	1.07	1.11	0.69
Materials, Supplies and Utilities			
Operating Supplies	0.02	0.02	0.04
Utilities	0.02	0.02	0.05
Maintenance Materials and Contract Labor	0.19	0.17	0.30
Catalysts and Chemicals	0.03	0.03	0.04
Total Materials, Supplies and Utilities	0.26	0.24	0.41
Labor			
Operating Labor and Supervision	0.07	0.06	0.11
Maintenance Labor and Supervision	0.09	0.09	0.14
Payroll Burden	0.06	0.06	0.10
Payroll Overhead/Administration	0.14	0.13	0.22
Total Labor Costs	0.36	0.34	0.57
G and A Overhead	0.03	0.03	0.04
Property Taxes and Insurance	0.22	0.20	0.36
Total Operating Costs Ratio	1.94	1.92	2.07

Table C-7 - Required Product Selling Price Ratios

Item	Relative Contribution to Unit Btu RPSP		
	Eastern Region	Appalachian Region	Western Region
Required Product Selling Price			
Capital	0.91	0.86	1.18
Coal	0.70	0.75	0.35
Operations	0.57	0.54	0.72
By-product Credit	-0.06	-0.06	-0.06
RPSP Ratios	2.12	2.09	2.19

Table C-8 - Possible and Required Annual Revenues and Benefit/Cost Ratio

Item	Relative Contribution to Annual Revenue		
	Eastern Region	Appalachian Region	Western Region
Possible Annual Revenue			
Products			
LPG, C_5	0.11	0.11	0.11
Gasoline	1.30	1.28	1.32
Product Total	1.41	1.39	1.43
By-products			
Sulfur	0.08	0.07	0.07
Sodium Sulfate	0.01	0.10	0.10
By-product Total	0.09	0.08	0.08
Total Annual Revenue Ratio	1.50	1.47	1.51
Required Annual Revenue Ratio (65/35 Debt/Equity at 12% DCFRR)	2.11	2.09	2.18
Benefit/Cost (B/C) Ratio	0.60	0.58	0.73

APPENDIX D
SINGLE PRODUCT SYNGAS BASED PROCESSES

APPENDIX D

SINGLE PRODUCT SYNGAS BASED PROCESSES

In order to supplement the information from the high potential processes, particularly the synthesis gas based indirect liquefaction type, cursory designs, estimates and economics were developed for plants producing SNG, using Fischer-Tropsch (FT) technology, and 95 percent hydrogen. These are briefly discussed below.

D.1 FISCHER-TROPSCH BASED SNG

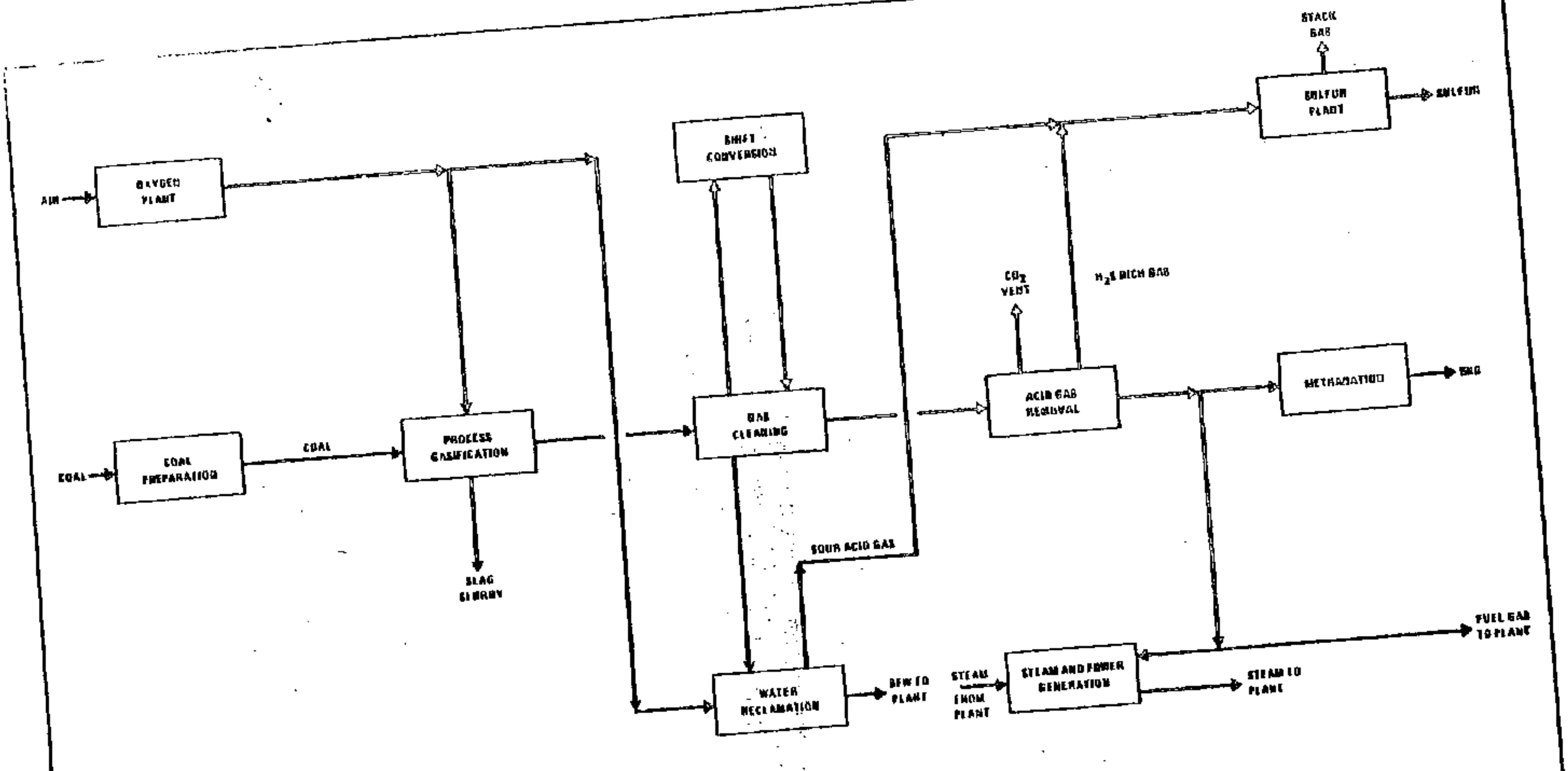
Inasmuch as the (FT) process configuration used here produced a high volume of SNG, a cursory modification of the FT process was made to produce all-SNG, having a total HHV of 600 billion Btu/day. Block flow diagram Drawing No. R-06-FS-1 depicts this process. Following gasification, shift conversion and methanation operations result in a SNG having an HHV of approximately 950 Btu/SCF. These units utilize the F-T flame sprayed catalytic reactors for consistent comparison with the F-T configuration used in the study.

The fixed capital investment estimate is approximately equal to that for the F-T plant included in this report as one of the high potential processes. In addition, the ratio of estimated possible total annual revenue to the required revenue to achieve a 12 percent discounted cash flow is 0.89 compared to 0.88 for the F-T SNG/liquid fuel process (SNG selling price of $3.25/MCF). At a SNG selling price of $4.25/MCF, the ratio becomes 1.15 compared to 1.01 for the all-SNG F-T plant and the SNG/liquid fuel plants, respectively. This, and other pertinent data, are presented in Tables D-1, D-2, D-3 and D-4.

D.2 HYDROGEN PROCESS

Another alternate single product process is presented for the manufacture of all hydrogen. From time to time, hydrogen has been touted as the desirable future fuel. A cursory process design was made to produce essentially 95 percent hydrogen, with approximately 4 percent methane, having a total HHV of 600 billion Btu/day. This process is shown on block flow diagram Drawing No. R-07-FS-1. Raw product from the gasifier (synthesis gas), following particulates removal and heat recovery, is shifted almost completely to hydrogen using conventional type shift reactors. The remaining small quantity of carbon monoxide is methanated in final clean-up catalytic methanation reactors.

The fixed capital investment estimate is 15 percent below that for the F-T based all-SNG plant. The ratio of estimated possible total annual revenue to the required revenue to achieve a 12 percent discounted cash flow is 0.97 compared to 0.89 for the all-SNG process. This is predicated on a hydrogen selling price equivalent to that of SNG on a Btu basis. This, along with other pertinent data, is summarized in Tables D-1, D-2, D-3 and D-4.



	TONS/DAY	MMSCFD	MMMBTU/D (HHV)
FEED COAL	35,030	—	880
SNG	14,580	635	600
SULFUR	1,184	—	—
TOTAL	15,764	—	600

THERMAL EFFICIENCY: $\frac{600}{880}$ X 100 = 68%

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COAL LIQUEFACTION PROCESS SURVEY
BLOCK FLOW DIAGRAM
SNG PLANT

THE RALPH M. PARSONS COMPANY
PASADENA, CALIFORNIA

JOB NO. 5658-1

DWG. NO. R-06-FS-1

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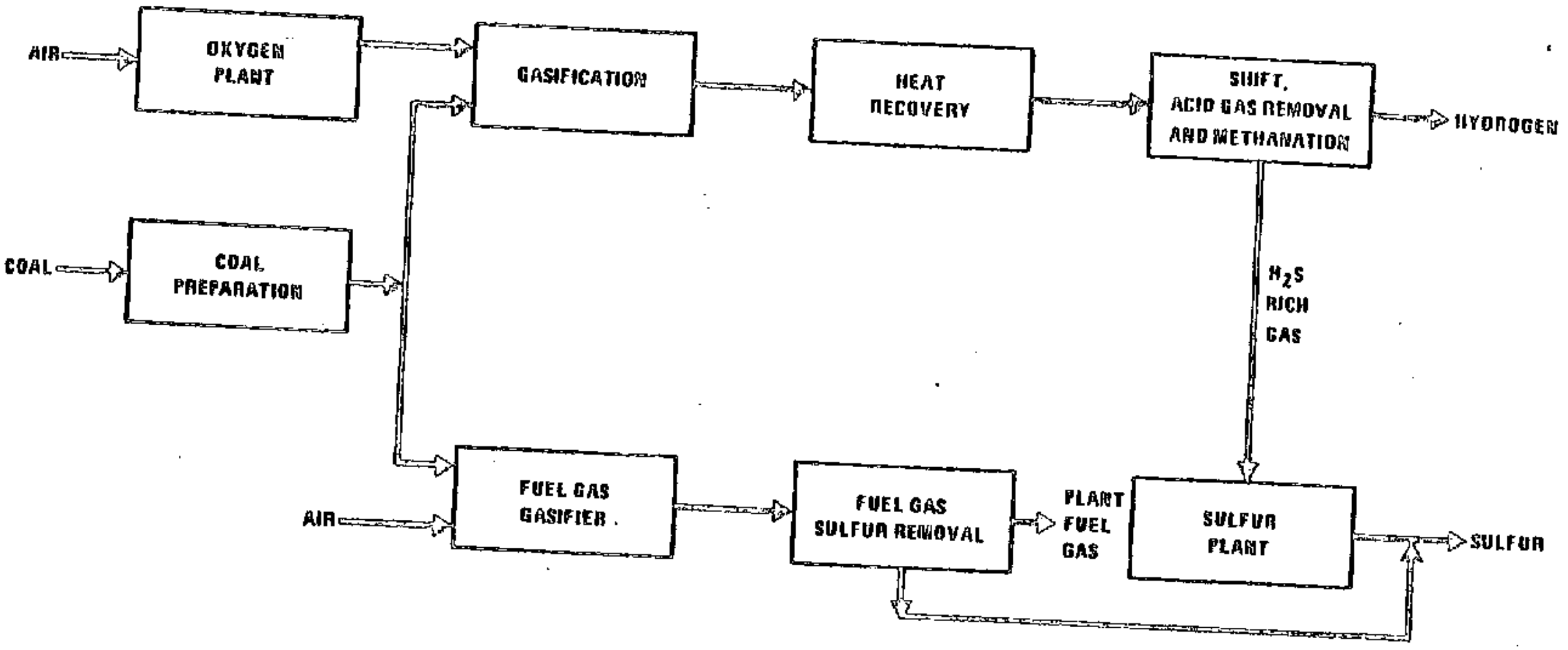

	TONS/DAY	MMSCFD	MMMBTU/D (HHV)
FEED COAL	35,980	—	900
HYDROGEN	6,390	1,719	600
SULFUR	1,189	—	—
TOTAL	7,579	—	608

THERMAL EFFICIENCY: $\frac{608}{900}$ X 100 = 67%

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COAL LIQUEFACTION PROCESS SURVEY
BLOCK FLOW DIAGRAM
HYDROGEN PLANT

THE RALPH M. PARSONS COMPANY PASADENA, CALIFORNIA	JOB NO. 5658-1	DWG. NO. R-07-FS-1	REV 0

Table D-1 - Ratios of Total Capital Investment SNG and H_2 vs. Fischer-Tropsch Eastern Region of Interior Coal Province (Base Case)

Item	Relative Contribution to Total Investment		
	Fischer-Tropsch Based	Fischer-Tropsch Based SNG	Hydrogen
Fixed Capital Investment	1.29	1.32	1.13
Initial Catalysts	0.02	0.02	-
Startup Costs	0.07	0.07	0.06
Interest During Construction	0.15	0.15	0.13
Commitment Fee	0.03	0.03	0.02
Total Depreciated Investment	1.56	1.59	1.34
Working Capital	0.12	0.12	0.10
Ratio to Total Capital Investment to SRC-II	1.68	1.71	1.44

Table D-2 - Sensistivity of Benefit/Cost Ratio to SNG Market Value Eastern Region of the Interior Coal Province (Base Case)

Item	Process		
	Fischer-Tropsch	F-T SNG	Hydrogen
Benefit/ Cost Ratio:			
@\$4.25/MCF SNG Price @\$1.52/MCF-H_2 Price[a]	1.01	1.15	1.25
@\$3.25/MCF SNG Price @\$1.16/MCF H_2 Price[a]	0.88	0.89	0.97
@\$2.25/MCF SNG Price @\$0.80/MCF H_2 Price[a]	0.74	0.63	0.69

[a] Hydrogen priced same as SNG on unit HHV

Table D-3 - Ratios of Operating Costs

Item	Relative Contribution to Operating Cost		
	Fischer-Tropsch based (%)	Fischer-Tropsch based SNG (%)	Hydrogen (%)
Coal	58.4	57.1	62.6
Materials, Supplies, and Utilities			
Operating Supplies	0.5	0.5	0.5
Utilities	0.5	0.5	0.7
Maintenance Materials and Contract Labor	10.4	10.6	9.5
Catalysts and Chemicals	2.1	3.6	1.0
Total Material, Supplies, and Utilities	13.5	15.2	11.7
Labor			
Operating Labor and Supervision	1.7	1.5	1.4
Maintenance Labor and Supervision	5.1	5.3	4.7
Payroll Burden	2.4	2.2	2.4
Plant Overhead/Administration	5.5	5.4	4.7
Total Labor Costs	14.7	14.2	13.2
G and A Overhead	1.5	1.5	1.5
Property Taxes and Insurance	11.9	12.0	11.0
Total Operating Costs	100.0	100.0	100.0
Operating Costs Ratio to SRC II	1.29	1.27	1.20

Table D-4 - Ratios of Unit Btu RPSP
SNG and H_2 vs. Fischer-Tropsch
Eastern Region of Interior Coal Province (Base Case)

Item	Relative Contribution to Unit Btu RPSP		
	Fischer-Tropsch based	F-T based SNG	Hydrogen
Capital[a]	0.63	0.66	0.50
Coal	0.49	0.47	0.49
Operations[b]	0.36	0.35	0.33
By-Product Credit[c]	-0.04	-0.04	-0.04
Ratio of RPSP to SRC II	1.44	1.44	1.28

[a] Depreciation; borrowing cost, 65% of the total investment borrowed at 9% interest, with principal repaid in equal installments over the 20 year project term; working capital, borrowed for the 20 year term; loan commitment fee of 0.75% on funds not yet drawn; 12% return on equity funds, income tax.

[b] Labor, utilities, maintenance labor and materials, property taxes, insurance, catalysts and chemicals, supplies, G&A overhead, plant overhead, payroll burden.

[c] By-product credit based on: sulfur, $60/ton; ammonia, $140/ton.

APPENDIX E

REFERENCE ECONOMIC DATA
CONVERSION OF RELATIVE ECONOMICS TO DOLLAR VALUES

Section 7 of this report presents relative economics for the four high potential processes. This method permitted achievement of the study objective which was comparison of processes.

To provide a basis for conversion of the relative economics to the first quarter 1978 preliminary estimates of dollar values Table E-1, titled "Key Reference Economic Data", is presented; this table presents dollar values for all key economic factors for the SRC II process which, based on these study results, was assigned a relative economic value of 1.0. The use of the E-1 economic information in conjunction with the Section 7 results will permit the reader to reproduce projected dollar values for the economics of any of the processes studied.

Table E-1 - Key Reference Economic Data
(SRC II Base Cases)

Item	Eastern Region Interior Coal Province	Appalachian Region, Eastern Coal Province	Western Rocky Mountain Region
Estimated Fixed Capital Investment ($MM)	1,155	1,132	1,270
Total Capital Investment ($MM)	1,475	1,450	1,580
Operating Costs ($/MM Btu)	1.75	1.90	1.15
Required Product Selling Price ($/MM Btu)	2.80	2.90	2.25
Possible Annual Revenues ($MM) (SNG Sold @ 3.25/MCF)	470	470	470
Required Annual Revenues ($MM) (65/35 Debt/Equity Ratio at 12% DCFRR)	555	575	445
Transportation Cost ($/MM Btu) New Athens, Ill. - Chicago, Ill.	0.12	-	-