LIQUEFACTION TECHNOLOGY ASSESSMENT

FINAL REPORT

work performed by
THE RALPH M. PARSONS COMPANY
Pasadena, California 91124
Subcontract No. 7186

MASTER

for the
FOSSIL ENERGY-DIVISION OF PROGRAM CONTROL AND SUPPORT
U. S. DEPARTMENT OF ENERGY

under subcontract with
OAK RIDGE NATIONAL LABORATORY
Oak Ridge, Tennessee 37830

operated by
UNION CARBIDE CORPORATION
NUCLEAR DIVISION

for the
U. S. DEPARTMENT OF ENERGY
Washington, D. C. 20545
Under Contract No. W-7405-eng-26

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LIQUEFACTION TECHNOLOGY ASSESSMENT

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ABBREVIATIONS

av	average
bbl	barrel(s)
B/C	benefit-to-cost ratio
BCR	Bituminous Coal Research
BNL	Brookhaven National Laboratory
BP	boiling point
BPD	barrels per day
BPCD	barrels per calendar day
BPSD	barrels per stream day
Btu	British thermal unit
Btu/D	British thermal units per day
Btu/lb	British thermal units per pound
C&C	catalysts and chemicals
CFFC	Clean Fuel From Coal
CSF	Consolidation Synthetic Fuel
DCF	discounted cash flow
D/E	debt/equity ratio
DOE	Department of Energy
eff	efficiency
EDS	Exxon Donor Solvent
EPRI	Electric Power Research Institute
ERDA	Energy Research and Development Administration
EUAC	equivalent uniform annual cost
FCI	fixed capital investment
FEA	Federal Energy Administration
FMC	Food Machinery Corporation
FSC	flame sprayed catalyst
F-T	Fischer-Tropsch
G&A	general and administrative
gpm	gallons per minute
H/C	hydrogen to carbon weight ratio
HHV	higher heating value
HRI	Hydrocarbon Research, Inc.
ID	Internal diameter
LPG	liquified petroleum gas
LTC	low temperature carbonization
M	thousand
MAF	moisture and ash free (coal)
MCF	thousand cubic feet

CONTENTS (Contd)

MM	million
MMM	billion
OCR	Office of Coal Research
OE	oil equivalent (6,000,000 Btu/bbl)
ORNL	Oak Ridge National Laboratory
PAD	Petroleum Administration for Defense
P&M	Pittsburg & Midway Coal Mining Co.
PDU	process development unit
PERC	Pittsburgh Energy Research Center
ppm	parts per million
ppmv	parts per million by volume
psi	pounds per square inch
psia	pounds per square inch absolute
psig	pounds per square inch guage
Quad	quadrillion, 10^{15}
R&D	research and development
RAR	required annual revenue
RMP	The Ralph M. Parsons Company
RPSP	required product selling price
scf	standard cubic feet
scfd	standard cubic feet per day
scfh	standard cubic feet per hour
sh	shell
SNG	substitute natural gas
sp gr and SG	specific gravity
SRC	Solvent Refined Coal
SRT	short residence time
syngas	synthesis gas
TLV	threshold limit value
TPD	tons per day
TPH	tons per hour
TPSD	tons per stream day
TPY	tons per year
UOP	Universal Oil Products
wt	weight

SECTION 1
INTRODUCTION

SECTION 1

INTRODUCTION

This report describes the results of an investigation of coal liquefaction processes that are under development. The main objectives of the investigation were to:

- Survey coal liquefaction processes being developed;
- Review the status of development;
- Make comparative technical and economic evaluations of high potential processes.

The survey was conducted under Subcontract No. 7186 with the Oak Ridge National Laboratory operated by Union Carbide Corporation, Nuclear Division (UCC-ND) under the Principal Contract No. W-7405-eng-26 with the Department of Energy (DOE), Fossil Energy Division of Program Control and Support. Their support and guidance are gratefully acknowledged.

The survey was a limited effort assignment relative to the broad scope of the objectives and the number of candidate coal liquefaction processes. It used the experience gained, previously and concurrently, in the development of seven comprehensive conceptual designs/economic evaluations plus reviews of design, construction and operation of several pilot plants. The conceptual designs included candidates from all generic liquefaction classifications, i.e., hydroliquefaction, donor solvent, pyrolysis and indirect.

Information used in the survey and evaluation was, in general, that available from published sources, provided by DOE, or obtained by nonproprietary consultation with process developers. Proprietary information is not included.

At the beginning of the survey more than 60 candidate processes were defined. From these candidates, a list of 33 were judged worthy of more extensive review. Approximately 30 processes were deleted from further consideration at this stage for various reasons. To illustrate, some were judged to have limited potential for commercialization, some appeared too similar to processes which were more advanced in their development cycle to warrant separate study, for some there was insufficient data available to us to permit detailed evaluation, and in some cases development work had been terminated by the developers.

The 33 processes reviewed represented an effective overview of the current state of coal liquefaction art. A summary of the characteristics of these processes was compiled and published; this "Data Source Book"¹ served as a starting point, as well as a reference, for the work reported here.

The survey results summarized in the following pages represent the equivalent of a "photograph," at a point in time (July, 1978), of judgements regarding the relative commercial potential for liquefaction processes. The judgements were based on analysis of information available to us within the limits of time and resources available.

SECTION 2
SUMMARY

SECTION 2

SUMMARY

A survey of coal liquefaction technology and analysis of projected relative performance of high potential candidates has been completed and the results are reported here.

The key objectives of the study included preparation of a broad survey of the status of liquefaction processes under development, selection of a limited number of high potential process candidates for further study, and an analysis of the relative commercial potential of these candidates.

Other objectives which contributed to successful achievement of the above key goals included definition of the characteristics and development status of known major liquefaction process candidates, development of standardized procedures for assessing technical, environmental, economic and product characteristics for the separate candidates, and development of procedures for selecting and comparing high potential processes. The comparisons were made for three production areas and four marketing areas of the U.S.

The survey was a limited effort in view of the broad scope of the objectives. It used the experience gained during preparation of seven comprehensive conceptual designs/economic evaluations plus comprehensive reviews of the designs, construction and operation of several pilot plants. Results and conclusions must be viewed in the perspective of the information that was available for use in the study, how this information was treated, and the full context of the economic comparison results.

Comparative economics are presented as ratios; they are not intended to be predictors of absolute values. Because the true cost of constructing and operating large coal conversion facilities will be known only after

commercialization, the relative values are considered to be more significant to achieve the objectives of this study.

There are many logic patterns that may be used for a technology assessment of this scope. As an overview prior to summarizing the results and conclusions, the particular logic pattern used here consisted of:

- Completion of a broad survey of coal liquefaction processes that are under development or have been operated commercially; more than 60 processes were defined.
- Selection of processes justifying further review; 33 of the 60-plus were selected.
- Development of a "Data Source Book" containing process characteristics and development status for the 33 processes; this served as a reference during the subsequent portions of the assessment program.
- Development of criteria for selection of high potential processes for more detailed study:
 - development status factors
 - technical factors
 - economic factors
 - other
- Selection of high potential processes; four were selected using the selection criteria described above plus other relevant criteria described later. Importantly, because of high interest, and potential for defined conditions, characteristics of one configuration of the Mobil-M process for production of gasoline are summarized here.
- Development of process configurations including predicted heat and material balances, product yields and thermal effi-

ciencies, environmental control procedures, product quantities/characteristics, and economics for the four high potential processes.

- Comparison of the economics for the four high potential processes:

- Fixed Capital Investment
- Total Capital Investment
- Operating Costs
- Required Product Selling Price (RPSP) at a 12% Discounted Cash Flow (DCF) profitability level for a defined project structure
- Possible product market values
- A benefit/cost (B/C) ratio, representing the ratio of possible market value to RPSP; this is a direct function of projected profitability.

- Retrospective Analysis of the results including:

- Comparison of relative economics developed for this study with results from comprehensive conceptual designs/economic evaluations prepared by major contractors.
- Economic impact of differences in product compositions, i.e., as more highly refined products are produced, what is the added cost?

- Development of conclusions and recommendations.

The contract required client agreement with the procedures used for technical and economic assessment; also agreement on the high potential processes selected for more detailed study.

The process configurations for all high potential candidates produced marketable products. Of key importance in a comparative assessment is to either make the comparisons based on virtually identical products or, alternatively, to adjust the comparison to quantitatively rationalize the economic impacts of differences in products generated by the candidate processes. In general, processes which produce heavy fuel products will require low capital investments and yield low annual revenues; the full profitability analysis envelope must be viewed to provide a meaningful comparison.

The information assembled for the 33 processes and published as a Data Source Book,¹ in conjunction with background from comprehensive conceptual designs/economic analyses and pilot plant performance reviews provided a portion of the basis for selection of high potential processes. Key elements of the selection criteria included:

- The development status; emphasis on the ability to permit commercialization by 1990.
- Capability of producing a product slate responsive to the historical energy consumption requirements for the producing area.
- Maximum profitability.
- Ecologically acceptable.
- A limited number of processes to be defined as "high potential" after assessing multiple technical, economic, product characteristics and development status factors; see Appendix A for details.

The preliminary assessment of 33 processes was completed using the above criteria guidelines. As a result, the following four high potential processes were selected:

- H-Coal based; a catalytic hydroliquefaction process.
- SRC II based; a pseudocatalytic hydroliquefaction process.
- Consolidation Synthetic Fuel (CSF) based; a donor solvent process.
- Fischer-Tropsch based using a flame sprayed catalyst system; an indirect process.

In conjunction with DOE, it was agreed that adequate detailed information was not available to us at the time (in 1977) to permit detailed objective inclusion of the Exxon Donor Solvent (EDS) process as a candidate. Additional information has subsequently become available and future assessments should include EDS.

Also, as indicated earlier, by virtue of special interest in the M-Gasoline and the high rating it obtained, the results of an independent evaluation of one configuration of the process is summarized in some detail in Appendix C. The information on this process available to us at that time contained what might be termed conservative technology for the gasification and methanol synthesis steps and therefore requires separate interpretation. Subsequently more detailed design and analysis would place it on a more comparable basis with the four high potential candidates.

A preliminary process package was developed for each of the candidates; this consisted of a block flow diagram, heat and material balances, product yields/characterizations, thermal efficiencies and environmental facility

design. Each plant had the capacity to produce fuels with an energy content of 600 billion Btu per day, equal to 100,000 barrels per day of fuel based on a nominal 6 million Btu per barrel reference value. The amount of feed coal required varied and was a function of the projected thermal efficiency.

Facilities were included in the design to upgrade and refine the coal liquids produced in the candidate process coal liquefaction step; product classifications included SNG, LPGs, gasoline, distillate fuel oils and residual fuel oils. These upgrading/refining additions caused us to introduce the terms "SRC II based," "H-Coal based," etc. to emphasize that while the basic coal liquefaction process to produce coal liquids was based on referenced data from the process developer, the added refining facilities were, in general, our design inputs.

The separate processes produced products in the SNG, LPG, gasoline, distillate fuel oil and residual fuel oil classifications, but they differed in quantities and characterization. Table 2-1 summarizes the product slates, feed coal quantity and other characteristics projected when processing Illinois No. 6 seam coal at an Eastern Region of the Interior Coal Province plant site; these results are considered typical of this Coal Province which include Kentucky, Indiana and Illinois. Table 2-1 shows: the SRC II based plant configuration used here would produce the greatest amount of residual fuel oil and least amount of gasoline; the H-Coal based plant would produce less resid and more gasoline; the CSF based plant would produce no residual fuel oil, a significant amount of distillate fuel oil, and about the same amount of gasoline as the H-Coal based plant; the Fischer-Tropsch based plant produced primarily gasoline and SNG with lesser amounts of LPG and distillate fuel oils. The SRC II based, H-Coal based and CSF based configurations all produce approximately the same amount of SNG, equivalent to 25-30% of total fuels produced on a Btu basis, while the Fischer-Tropsch based plant produces approximately twice as much SNG. Therefore, while the configurations used, in general, produced fuels in the target product categories, they tend to be lighter products and more highly refined as you move from left to right in Table 2-1. This should be kept in mind as the summary of relative economics unfolds.

Estimates of capital investment requirements, operating costs, and RPSP at 12% DCF were prepared for each candidate. All economic factors were developed based on first quarter 1978 dollars. The fixed capital investment (FCI) estimates were based on an estimate specific to the primary coal liquefaction operations as defined by non-proprietary information provided by the process developer plus estimates of coal preparation and coal derived liquids/gas recovery and refining based on prior comprehensive conceptual design results; also experience factors for upgrading/refining operations from the petroleum refining industry. More details of the estimating procedures are presented in Section 7. Operating costs were developed by an analysis of operating labor, catalyst and chemicals, utilities and other requirements; in general, all plants captively produced their steam and electricity needs. The RPSP indicates the equivalent uniform annual revenue required to generate a 12% DCF rate of return, first quarter 1978 dollars, using a 65/35 debt/equity project structure, 9% interest rate, 5-year construction period and 20-year operating period.

The possible product market values were next developed for each candidate process by comparing projected coal derived product characteristics with current petroleum-based product characteristics and market values. In the case of SNG, a judgemental value of \$3.25 was selected for the base case analysis; sensitivity of relative economics to SNG market value over the range of \$2.25 to \$4.25 was also developed. The \$3.25 value is intermediate between natural gas pricing and projected cost of SNG from coal and the allowed pricing for incremental SNG from naphtha.

The marketability and market values of coal derived liquid and gaseous fuels requires confirmation. This should be an important goal of the national coal conversion development program.

Finally, a B/C was developed which represents the ratio of possible annual revenue to RPSP at 12% DCF. This ratio is significant since the higher the value the higher the indicated potential profitability.

The results of the comparative economic analysis show the following rankings, where a ranking of 1 represents the most favorable economics and higher numbers successively less favorable.

<u>Process</u>	<u>Benefit/ Cost Ratio (B/C)</u>	<u>Required Product Selling Price @ 12% DCF (RPSP)</u>	<u>Possible Annual Revenue</u>	<u>Capital Investment Requirements</u>	<u>Operating Costs</u>
SRC II Based	1	1	4	1	1
H-Coal Based	1	2	2	2	2
CSF Based	2	3	3	3	3
Fischer-Tropsch Based	1	4	1	4	4

These results indicate that, proceeding from the SRC II based configuration to Fischer-Tropsch in the above table:

- The products tend to become lighter and more highly refined.
- The capital investment and operating costs increase.
- The possible annual revenues and RPSP increase; the inversion of H-Coal based and CSF based is an exception to this generalization.
- The B/C ratio, an indicator of potential profitability, groups SRC II based, H-Coal Based and Fischer-Tropsch based technologies as equivalent, with CSF based having the lowest indicated B/C ratio.

Analysis in retrospect indicates that the Fischer-Tropsch configuration used for this study has the greatest potential for relative improvement among the candidates listed.

The economics of the Fischer-Tropsch based configuration used is most sensitive to SNG market value. At \$2.25/MCF, its B/C ratio is lower than H-Coal and SRC II based, at \$3.25 the three are equivalent, and at \$4.25, Fischer-Tropsch based clearly shows the highest B/C (profitability).

Recognition of the critical effect of product composition on the comparative economics generated the objective to develop a correlation between product compositions and key economic parameters such as fixed capital investment, product market value and required product selling price. The results, considered significant, indicate that each of these parameters correlate well with the hydrogen to carbon (H/C) weight ratio of the products. The logic is that coal liquefaction essentially involves adding hydrogen to coal; coal has a H/C ratio of about 0.07, aromatic liquids such as benzene a ratio of approximately 0.08, aliphatic liquids such as heptane about 0.19, and SNG (methane) about 0.33. The addition of the hydrogen consumes added material such as coal and water plus requires additional equipment.

The correlations of H/C vis-a-vis economic parameters presented in Figures 8-1, 8-2, 8-3, and 8-4 in this report provide a unique contribution to the field of coal liquefaction as a direct result of our contractual work.

The H/C correlations also indicate that there is only a slight difference in the economics of the Fischer-Tropsch based plant and a SNG plant using similar gasification/gas purification procedures. This may be interpreted as confirmation that the costs of producing a purified mixture of carbon monoxide and hydrogen (syngas) dominates the economics of indirect liquefaction processes such as Fischer-Tropsch.

The process designs and economic evaluations developed here were allotted modest effort relative to that applied to more comprehensive conceptual designs by major contractors. In order to shed some light on the question of whether differing conclusions would have been reached if significantly more effort were applied, the results of conceptual designs/economic evaluations for SRC II, H-Coal, CSF and Fischer-Tropsch were compared. In this case, the SRC II based and Fischer-Tropsch based designs/economic evalua-

tions were developed by one major contractor while the H-Coal based and CSF based units were developed by a second major contractor. The trends between designs/economics for the processes, when generated by the same contractor, were similar to those developed in this assessment.

Assessments and comparisons were also developed for two additional production plant sites and four marketing areas. Other production locations were the Appalachian Region of the Eastern Coal Province and the Powder River Region of the Rocky Mountain Province; Pittsburgh No. 8 seam coal and a sub-bituminous coal typical of a Wyoming location were used for feeds to these plants, respectively. To provide a basis for estimating transportation costs, the production plants were arbitrarily sited at New Athens, Illinois; Moundsville, West Virginia and Gillette, Wyoming; some site data was available for each of these locations based on the results of earlier studies. The marketing locations, defined by Oak Ridge National Laboratory, were Delaware, Chicago, Houston and San Francisco.

The most detailed work was done on the plant conceived to be located in Illinois; this is termed the "base case." The base case was then modified for the other two locations based primarily on differences in feed coal composition. The results indicate that the FCIs tend to be slightly lower in the Appalachian Region and 3-15% higher in the Rocky Mountain Region when compared with the base case. The RPSPs were similar for the Appalachian and base cases but were higher for the Rocky Mountain Region. Similar results were indicated for the B/C ratios.

Estimates were developed for transportation costs from the production sites to the indicated marketing areas; for the base case, these amounted to 3-6% of the plant F.O.B. RPSP with the lowest being for CSF and the highest Fischer-Tropsch. In general, transportation costs from the Appalachian Region to Dover, Delaware were approximately twice as great as the Interior Region and for Gillette, Wyoming to San Francisco, two-to-four times as great.

This study has looked closely at high potential candidates which are capable of supplying synfuels by 1990. The importance of development of a viable

coal conversion capability in the U.S. requires that the analysis effort exemplified by this survey be continued and updated as new information becomes available.

As improvements are made in the liquefaction technologies, there is an incentive to maintain an active cross-fertilization program to minimize the time required to incorporate the improvements and/or problem solutions generated by one technology to other technologies also under development.

Details of the survey and assessment are contained in the following sections of this report.

Table 2-1 - Coal Feed and Product Summary
Base Case: Eastern Region of Interior Coal Province

Item	SRC II Based			H-Coal Based			CSF Based			F-T Based		
	TPI	BPSD (10 ⁶ SCFD)	10 ⁹ Btu/D	TPI	BPSD (10 ⁶ SCFD)	10 ⁹ Btu/D	TPI	BPSD (10 ⁶ SCFD)	10 ⁹ Btu/D	TPI	BPSD (10 ⁶ SCFD)	10 ⁹ Btu/D
<u>Feed</u>												
Coal	30,627		767.9	32,257		808.8	35,714		896.4	36,341		912.1
<u>Products</u>												
SNG	3,422	(140)	155.1	3,703	(155)	165.6	3,515	(135)	151.6	7,274	(325)	316.0
C ₃ -LPG	452	5,101	19.6	422	4,753	18.3	1,040	11,723	44.7			
C ₄ -LPG	218	2,133	9.3	114	1,117	4.9	278	2,476	11.9	1,186	11,751	50.7
Gasoline	1,579	10,337	55.5	2,941	21,965	118.1	2,822	20,973	113.6	4,643	37,536	197.0
Oxygenates										642	3,759	13.6
Distillate Fuel Oil No. 2										413	2,723	15.9
Distillate Fuel Oil No. 4				481	3,005	18.6	4,870	46,983	278.2			
Distillate Heavy Fuel Oil							3,219			168	1,076	6.8
Residual Fuel Oil	10,482	52,309	360.6	7,725	40,151	274.5						
<u>By-Products</u>												
Sulfur	987		7.9	1,074		8.6	1,201		9.6	1,228		9.8
NH ₃	97		1.9	128		2.5	182		3.6			
<u>Product Ratios</u>												
Total Fuel Products												
MM Btu/D		600			600			600			600	
MM Btu/ton of coal		19.6			18.6			16.8			16.5	
Liquid Products												
bbl/D		69,880			71,030			82,155			56,845	
bbl/ton of coal		2.3			2.2			2.3			1.6	

SECTION 3
SELECTION OF HIGH POTENTIAL PROCESSES

SECTION 3

SELECTION OF HIGH POTENTIAL PROCESSES

A prime objective of this study was to select a limited number of high potential coal liquefaction processes and then develop documented opinions regarding their rankings with regard to potential technical performance, pertinent economic factors, and product marketability. The selection of the high potential processes was accomplished by establishing selection criteria plus evaluation procedures to display the projected performance of the separate processes in comparison with the selection criteria. This section will describe the logic pattern used to select four high potential processes for more detailed analysis.

3.1 SELECTION PROCEDURES

Prior work under this contract had defined sixty-plus liquefaction process candidates and had reported the characteristics and development status of thirty-three of what were considered the most promising of the processes.¹ The processes were arbitrarily classified and given process numbers as follows:

1. Hydroliquefaction

1.1 Direct, Non Catalytic

1.1.1 Solvent Refined Coal (SRC-I)

1.1.2 Solvent Refined Coal (SRC-II) (Pseudocatalytic)

1.1.3 UOP Extraction

1.2 Direct, Catalytic

1.2.1 Gulf Catalytic Coal Liquids (CCL)

1.2.2 SYNTHOIL

1.2.3 H-Coal

1.2.4 Clean Fuel From Coal (CFFC)

1.2.5 CONOCO Zinc Halide Hydrocracking

1.3 Indirect, Donor Solvent

1.3.1 Exxon

1.3.2 ADL Extractive Coking

1.3.3 Consolidation Synthetic Fuel (CSF)

2. Pyrolysis

2.1 Direct

2.1.1 Char-Oil-Energy Development (COED-RMP)

2.1.2 Char-Oil-Energy Development (COED-FMC)

2.1.3 Occidental Research Corp. Coal Flash Pyrolysis

2.1.4 TOSCOAL

2.1.5 Lurgi-Ruhrgas

2.2 Hydrocarbonization

2.2.1 U.S. Steel Clean Coke

2.2.2 Coalcon

2.3 Rapid Hydrogenation

2.3.1 BNL Rotating Fluidized Bed

2.3.2 Short Residue Time (SRT) Hydropyrolysis

2.3.3 Intermediate Coal Hydrogenation

2.3.4 Schroeder's Rapid Hydrogenation

2.3.5 BNL Flash Hydropyrolysis

2.3.6 Rockwell Direct Coal Hydrogenation

3. Indirect

3.1 Fischer-Tropsch

3.1.1 Flame Sprayed Catalyst

3.1.2 ARGE

3.1.3 SYNTHOL

3.2 Indirect, Methanol

3.2.1 High Pressure Methanol

3.2.2 ICI LP/LT Methanol

3.2.3 Lurgi LP/LT Methanol

3.2.4 Three Phase Methanol Synthesis

3.3 Indirect, Other

3.3.1 M-Gasoline (Mobil)

4. Miscellaneous

4.1.1 Supercritical Gas Extraction (SGE)

4.1.2 British National Coal Board

Principle categories of this particular classification system are hydroliquefaction, pyrolysis and indirect liquefaction, with donor solvent processing classified as a subdivision of hydroliquefaction.

The procedure used to achieve the objective of selecting high potential processes from this list of thirty-three candidates is described in the following paragraphs.

A list of selection criteria elements is shown in Table 3-1. Key elements include:

- The successful candidates must be capable of commercialization by 1990.
- The processes are to be tested for their ability to produce a product slate typical of the fuel consumption patterns for each geographical area as shown in Table 3-2 using feed coal to the process plants with typical analyses as summarized in Table 3-3.
- High profitability and ecologically satisfactory.

- Definition of the candidate as a high potential process by assessment of multiple technical, economic, product characteristic and development status factors.

To accomplish the broad range technical/economic/status assessment described in the last criteria listed above, a detailed procedure was developed and implemented; details of the techniques used and the results are presented in Appendix A.

3.2 HIGH POTENTIAL CANDIDATES

Using the above criteria, the selection process was completed and the following four processes were chosen as high potential processes:

- SRC II; a pseudocatalytic hydroliquefaction process
- H-Coal; a catalytic hydroliquefaction process
- CSF; a donor solvent process
- FSC (Fischer-Tropsch using Flame Sprayed Catalyst); an indirect process to produce hydrocarbons.

The M-Gasoline process is not included in the above list of candidates but was however, evaluated because of special interest in the process and the high rating it obtained. The evaluation basis, established for another study, differed from that adapted for the above candidate list. The Mobil-M results are therefore presented separately for reference in Appendix C of this report.

The Exxon Donor Solvent process (EDS) was not included in the more detailed analyses because, as decided mutually with DOE, sufficient information was not available to us at the beginning of our work.

Table 3-1 - Selection Criteria for High Potential Processes

Criteria	Comment
1. Time Frame	Starting a commercial design in 1985 and start up in 1990
2. Product Slate	A product slate is selected based on product and energy profiles for each geographical area shown in Table 3-2
3. Maximum Profitability	—
4. Ecological Acceptability	—

Table 3-2 - Target Product Slate By Geographical Area

Item	Appalachian Region of Eastern Coal Province		Eastern Region of Interior Coal Province		Powder River Region of Rocky Mountain Coal Province	
	Btu/day (Billion)	(%)	Btu/day (Billion)	(%)	Btu/day (Billion)	(%)
SNG	171	28.5	269	44.8	202	33.7
LPG	15	2.5	11	1.8	13	2.2
Gasoline	106	17.7	182	30.3	183	30.4
Distillate Fuel Oil	125	20.9	106	17.7	111	18.5
Residual Fuel Oil	183	30.4	32	5.4	91	15.2
	—	—	—	—	—	—
Total	600	100.0	600	100.0	600	100.0

Table 3-3 - Typical Feed Coal Analyses^a By Coal Source Area

Item	Appalachian Region of Eastern Coal Province	Eastern Region of Interior Coal Province	Powder River Region of Rocky Mountain Coal Province ^b
Proximate Analysis wt (%)			
Volatile Matter	40.5	38.5	44.2
Fixed Carbon	49.5	51.7	45.1
Ash	7.3	7.1	8.0
Moisture	2.7	2.7	2.7
	<hr/>	<hr/>	<hr/>
Total	100.0	100.0	100.0
Ultimate Analysis wt (%)			
Carbon	71.6	70.7	66.3
Hydrogen	5.0	4.7	4.7
Oxygen	7.7	10.3	16.5
Sulfur	4.4	3.4	0.8
Nitrogen	1.3	1.1	1.0
Moisture	2.7	2.7	2.7
Ash	7.3	7.1	8.0
	<hr/>	<hr/>	<hr/>
Total	100.0	100.0	100.0
Gross Heating Value			
Btu/lb	12,640	12,550	11,211
^a This is feed coal to Process Plant, cleaned and dried as necessary. ^b Run of mine subbituminous coal from this area contains approximately 25% water and has a gross heating value of about 8,720 Btu/lb.			

SECTION 4
GEOGRAPHICAL AREA CONSIDERATIONS

SECTION 4

GEOGRAPHICAL AREA CONSIDERATIONS

An objective of the study was to assess the projected relative performance/economics of the candidate processes in multiple areas of the country while supplying the local fuel markets. The goal was also to assess whether one process might be preferable in one geographical area while an alternate process would be preferable in another area.

To respond to these objectives, the study considered three production areas, three typical feed coal compositions, and four marketing areas; these are summarized in Table 4-1.

The synfuels plant designs/economics were based on plant sites in the coal producing regions using purchased clean washed coal to produce, in a manner environmentally acceptable to the area, a product slate responsive to the area's historical consumption pattern. Relative F.O.B. plant gate economics were developed for the separate processes. In addition, planning type estimates of transportation costs were developed for movement of the products to the four market locations listed in Table 4-1; "delivered cost" relative economics were then also developed.

In recognition of the practical restraints of time and resource allocation available for this study, a conceptual process design and economic evaluation was developed in most detail for the Eastern Region of the Interior Coal Province location — referred to in this report as the Base Case. Subsequently, adjustments to the Base Case were made to project process performance and economics for the other two production locations; experience developed during prior preparation of designs for multiple locations used.²

4.1 PRODUCT/MARKET SLATES

Target product market slates were selected based on historic fuels consumption profiles for each marketing location identified in Table 4-1. Profiles for these areas were developed using 1977 Monthly Petroleum Statistics Reports published by the Federal Energy Administration (FEA)/National Energy Information Center. Product supply/demand statistics in these reports are given by Petroleum Administration for Defense (PAD) Districts. Districts I, II, III, IV, and V were grouped as follows to approximate the locations listed in Table 4-1:

<u>Defense Districts</u>	<u>States</u>
PAD I - Eastern	All of New England, New York, Pennsylvania, New Jersey, Maryland, Delaware, West Virginia, Virginia, North Carolina, South Carolina, Georgia, and Florida
PAD II & III - Central	North Dakota, South Dakota, Nebraska, Kansas, Oklahoma, Minnesota, Iowa, Missouri, Wisconsin, Illinois, Michigan, Indiana, Kentucky, Tennessee, Ohio, New Mexico, Texas, Arkansas, Louisiana, Mississippi, and Alabama
PAD IV & V - Western	Montana, Idaho, Wyoming, Utah, Colorado, Oregon, Washington, California, Nevada, Alaska, and Hawaii

Petroleum products reported by FEA include motor gasoline, aviation gasoline, naphtha-type jet fuel, kerosene-type jet fuel, kerosene, distillate fuel oil, No. 4 fuel oil, and residual fuel oil. For the purpose of this study, these liquids were realigned to produce LPGs gasoline, distillate fuel oils and residual fuel oil.

As a next step, energy consumption profiles were developed for each geographical area based on published statistics.^{3,4} This provided the consumption ratios of natural gas to petroleum liquids by area. The target product slate as summarized in Table 3-2 was then developed using this ratio as a guideline.

Development of the Table 3-2 target product slates defined the specific products to be made and the level of coal derived liquid upgrading required to produce these products. It permitted the designers/evaluators to develop the process configurations presented in Section 5 of this report.

Table 4-1 - Geographical Factors

Production Location			Typical Feed Coal Composition		Marketing Location
Coal Region	Coal Province	Typical Site	Type	Reference	
Appalachian	Eastern	Moundsville, West Virginia	High Volatile Bituminous A Pittsburgh No. 8	Table 3-2	Delaware
Eastern	Interior	New Athens, Illinois	High Volatile Bituminous B Illinois No. 6	Table 3-2	Chicago Houston (secondary)
Powder River	Rock Mountain	Gillette, Wyoming	Subbituminous	Table 3-2	San Francisco

SECTION 5
DESCRIPTIONS OF HIGH POTENTIAL PROCESSES

SECTION 5

DESCRIPTION OF THE HIGH POTENTIAL PROCESSES

This section contains development status summaries, descriptions, operating conditions and estimated product yields for the four selected high potential processes; block flow diagrams are included. The sources of process data and cost information for each of the processes are also summarized.

Similar information for Mobil-M technology is contained in Appendix C.

Each process design is based on a production capacity of fuel products having a total HHV of 600 billion Btu/day; products include SNG, LPG, gasoline, distillate fuels and residual fuels. To increase the uniformity of product slates from the separate processes and provide as common a basis of comparison as possible, some process modifications and additions were incorporated. An example is the addition of hydrotreating, catalytic reforming, polymerization, additional LPG recovery and distillation units to the Fischer-Tropsch processes; these process steps convert the naphthas to gasoline product. An earlier design⁵ produced light and heavy naphtha as part of the product slate.

The SRC II process, included in the high potential process group, is based on the results of Run 11 as presented in P&M's October 1977 monthly report. The configuration used is similar to Parsons Oil/Gas conceptual design.⁶ Thus, the solids removal filtration step is included in the process. The process is of the SRC II mode wherein a slurry solvent is recycled to the dissolver as distinguished from the SRC I use of a solids-free distilled recycle solvent to slurry the feed coal. Subsequent to the beginning of this study, development efforts indicated elimination of the filtration step to be a practical improvement.^{7,8} Since the development was not yet sufficiently defined at that time, the filtration mode was retained.

Assessments of eastern and western plant locations operating on Eastern Coal Province coal, and Rocky Mountain Coal Province coal, respectively, are also included for each of the high potential processes. The difference in coal compositions result in processing and cost variations from the base case operation. The three cleaned and dried feed coals' analyses involved were presented in Table 3-3.

5.1 SRC II BASED PROCESS

5.1.1 DEVELOPMENT STATUS

The SRC process has been under development using government funds in the U.S.A. for about 15 years. There are two pilot plants in operation; a nominal 50-ton-per-day DOE-sponsored unit at Tacoma, Washington and a 6-ton-per-day DOE/EPRI/Southern Services Corporation sponsored unit at Wilsonville, Alabama. The Tacoma pilot plant was designed to process approximately 50 short tons per day of coal when practiced in the SRC I mode. SRC I, in general, produces a solid product and removes ash and unreacted coal by filtration prior to product recovery. Depending on the composition of the feed coal and processing conditions, the SRC I may have a sulfur content of the order of 0.8% and contain about 0.15% solids. It is intended for boiler fuel use.

The Tacoma pilot plant has also been operated in the SRC II mode; in this mode a portion of the unfiltered reactor product is used to slurry feed coal. The result is a higher ash content and longer effective reaction time in the coal dissolver, causing a higher percentage of hydrogen addition to the coal and production of products which are more liquid than for the SRC I mode of operation.

The SRC pilot plant was mechanically completed in late 1974 and has operated since that time. It has successfully produced 3,000 tons of SRC I product which was then used in a boiler performance test to determine the functional performance of the product. Subsequently, the pilot plant has been operated in the SRC II mode successfully to obtain operating

data and also to produce 5,000 barrels of liquids for product testing. It has had an onstream performance record of 85-plus percent in the SRC II mode vis-a-vis about 68 percent in the SRC I mode.⁸

The Tacoma pilot plant is now being returned to SRC I service. During future SRC I operations, new designs for equipment, including filters, will be tested. The Wilsonville pilot plant has operated successfully while testing a number of coal types in the SRC I mode.

5.1.2 DATA SOURCES AND FIXED CAPITAL INVESTMENT BASES

The data for the SRC II coal liquefaction process was obtained from reports submitted to the Department of Energy and its predecessor, The Energy Research and Development Administration, by Pittsburgh and Midway Coal Mining Company (P&M) under Contract No. EX-76-C-01-496. The design yields are based on a correlation of yields of gas and liquid products from MAF coal versus hydrogen consumption on MAF coal developed from eleven SRC II material balance runs, performed at the Tacoma Pilot Plant in 1977 (runs SR-1 through SR-11).⁸ The dissolver operating conditions selected were those used for run SR-11. Pilot plant operations for this run included filtration. Subsequent runs have been made eliminating the filtration step.

Basic data and yields for the gasification unit originated from work done for the Office of Coal Research (OCR) by Bituminous Coal Research, Inc. (BCR) under Contract No. 14-32-0001-1207.⁹ An independent design was developed by Parsons for the specific conditions of temperature, pressure and gas composition required for the SRC II process plant.

The basic information for the fixed capital investment estimate of the units shown in the block flow diagram R-01-FS-1 were obtained from:

1. A report prepared by Parsons for ERDA under Contract No. E(49-18)-1775 published in March 1977, entitled "Oil/Gas Complex-Conceptual Design/Economic Analysis", Report No. 114 - Interim Report No. 4.⁶

2. A report prepared by Parsons for DOE under contract No. EX-76-C-01-1775 published in December 1977, entitled "Project POGO - Coal Refinery Complex-Conceptual Design/Economic Analysis-Power-Oil-Gas-Other Products", R&D Report No. 114 - Interim Report No. 6.²
3. Parsons in-house information.

5.1.3 PROCESS DESCRIPTION

The SRC Process, as shown in the block flow diagram, Drawing R-01-FS-1, contains the following coal conversion units:

- A direct, pseudocatalytic hydroliquefaction unit (Coal Slurrying and Coal Dissolving and Product Let-Down) to convert coal to the primary products: SNG, LPG, naphtha and fuel oil;
- A two-stage, entrained-bed, slagging Process Gasifier to convert coal to methane, syngas and minor amounts of byproducts;
- A two-stage, entrained-bed, slagging Fuel Gas Gasifier to convert coal and filter cake to plant fuel gas.

In this process system, 95%-minus-20-mesh coal is combined with a filtered recycle solvent and unfiltered slurry recycle and fed to a coal slurrying unit. The coal slurry is continually pumped through the preheat furnace. Make-up plus recycle hydrogen is added at the furnace entrance; the resulting mixture is preheated, and then fed to the dissolver, which is operated at an outlet temperature of about 850°F and 2,000 psig. The temperature rise in the dissolver is controlled with a hydrogen-rich quench stream from the product let-down system.

The product mixture from the dissolver consists of a gas phase, a liquid phase and a solid phase.

The gas phase is separated in the product let-down system, treated to remove CO_2 and H_2S in the acid gas removal unit, and sent to drying and cryogenic separation. A hydrogen-rich gas is separated cryogenically, compressed and recycled to the dissolver. The excess hydrogen and syngas is methanated to a methane-rich gas which is purified and compressed to produce pipeline quality SNG. The condensate from the Drying and Cryogenic Separation unit are fractionated to produce C_3 and C_4 LPG.

The liquefied coal products, containing undissolved coal and ash, are subjected to fractionation in the atmospheric distillation unit. The unit produces the following products:

1. Naphtha (C_5 to 400°F boiling range).
2. Light distillate (400°F to 500°F boiling range). Used as a constituent of the fuel oil products.
3. Heavy distillate (500°F to 650°F boiling range) used both as wash oil in the filters and as a constituent of the fuel oil product.
4. Atmospheric bottoms (650°F+) containing the solid residue consisting of undissolved coal and ash; used both as process solvent (slurry recycle) and filter feedstock.

The fractionated naphtha is further hydrotreated and catalytically reformed to produce a high octane gasoline. A portion of the 650°F-plus liquids produced are combined with the net 400°-650°F liquids produced in the atmospheric fractioner to make a low-sulfur residual fuel oil.

The solids in the atmospheric bottoms (650°F-plus) consist of undissolved coal and ash and are recovered in the form of a filter cake in the filtration unit. The filter cake is dried to recover the wash solvent. The heating value of the dry filter cake is recovered by combining it with coal and gasifying it to produce plant fuel gas.

The plant is a self-contained facility producing all of its required electrical power, steam and fuel gas; the only import utility required is water.

5.1.4 ASSESSMENT OF PROCESSING OTHER COALS

Table 5-1 is presented here to compare SRC II yields for the Interior Coal Province (Illinois No. 6 Seam Coal) and the Eastern Coal Province (Pittsburgh Coal). The data in the table was developed using bench-scale equipment by P&M at their Merriam, Kansas laboratory⁸ and is illustrated here for comparison purposes only. The data for the SRC II Case as shown in the block flow diagram originated from the Tacoma pilot plant as previously referenced and not from the Merriam laboratory data shown in Table 5.1.

Since the operating conditions for the Run GU-182R (Illinois No. 6 coal) and Run GU-181R (Pittsburgh seam coals) are almost identical, comparison between the two coals can be made.

The results indicate that Pittsburgh seam coal is at least equivalent to Illinois No. 6 coal as a feedstock for SRC II. This is evidenced by the following:

1. Pittsburgh coal has a higher total distillate (38.7% vs. 34.3%).
2. Pittsburgh coal has an equivalent MAF conversion (93.21% vs. 92.99%).
3. The Pittsburgh coal has a lower SRC, 850°F-plus, yield with more distillate liquids in the Residual Fuel Oil mix.

P&M reported a high iron content (2.5 weight per cent) for Pittsburgh coal, suggesting that this could be a factor in its high reactivity.

Yield data for SRC II processing of Rocky Mountain coal is sparse. Projected yields for processing Wyodak coal are presented in a report titled "Project POGO-Coal Refinery Complex-Conceptual Design/Economic Analysis," prepared by Parsons for DOE under contract No. EX-76-C-01-1775.² In this report, the yields for SRC I mode of operation on Rocky Mountain Coal were adjusted to reflect SRC II operation. The results indicate dissolver oil yields similar to those obtained from Illinois No. 6 coal when operating conditions are identical; however, MAF conversion for the Rocky Mountain coal is low compared to other coals.

Since it has a relatively high oxygen content, the hydrogen consumption required to remove the oxygen and produce water is expected to be high. Another inherent disadvantage to processing Rocky Mountain coal would be the high run-of-mine water content, approximately 25%, which would require extensive drying facilities. The net effect of processing the Rocky Mountain coal would be to lower the thermal efficiency compared to the Pittsburgh and Illinois No. 6 coal. This is primarily due to:

1. The higher drying load due to the high moisture content.
2. Additional hydrogen required to convert the oxygen to water.

This would result in a higher fixed capital investment for the same quantity of fuel products in a commercially sized plant for the Rocky Mountain coal. However, the higher investment would tend to be compensated for by the significantly lower cost of Rocky Mountain coal when compared to the Appalachian and the Interior coals.

Table 5-1 - SRC II Laboratory Data For Several Coals^a

Item	Interior Coal Province (Illinois No. 6 Seam)	Eastern Coal Province (Pittsburgh Seam, Ireland Mine)
Merrian Lab Run No.	GU-182R	GU-181R
Yields: (Weight % Dry Coal)		
H ₂ O	8.2	5.7
Gas:		
C ₁ -C ₄	10.8	12.9
Other Gases	4.3	4.6
Light Distillate	9.9	13.2
Heavy Distillate	24.4	25.5
SRC	27.4	22.1
IOM (Insoluble Organic Matter)	6.1	5.9
Ash	12.3	13.9
Total	103.4	103.8
H ₂ Reacted (%)	3.4	3.8
MAF Conversion(%)	92.99	93.21
Total Gas Yield (%)	15.1	17.5
Total Distillate Oil Yield (%) (Light + Heavy Distillate)	34.3	38.7
Total Distillate Oil + SRC Yield (%)	61.7	60.8
Residual Fuel Oil (%) (Heavy Distillate + SRC)	51.8	47.6
Distillate Liquids in Residual Fuel Oil Mix (%)	47	54
Conditions		
Pressure (psig)	1900	1900
Nominal Dissolver Temperature (°F)	851	851
Nominal Residence Time (Hr)	0.98	0.99
Coal Space Rate (Lb/Hr/Ft ³)	22.0	21.8
Slurry Composition		
Coal (%)	30	30
Unfiltered Recycled Dissolver Product (%)	70	70

^aP&M August 1977 Monthly Progress Report for ERDA Under
Contract No. EX-76-C-01-496

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5.2 H-COAL BASED PROCESS

5.2.1 DEVELOPMENT STATUS

The H-Coal process has been under development by Hydrocarbon Research, Inc. (HRI), since the mid-1960s with funding from both private sector and public sector sources. Test facilities, consisting of a 25-lb/day bench-scale unit, and a 3 TPD process development unit (PDU) have been operated at Trenton, New Jersey. Construction is now in progress for a major pilot plant located at the Catlettsburg, Kentucky Ashland Oil Refinery, with construction scheduled to be complete in 1979. The design capacity of this pilot plant is 250 tons per day of feed coal when operated to produce syncrude and as much as 600 tons per day when operated to produce fuel oil. The project costs are projected to be in excess of \$200 million U.S.A. dollars.

5.2.2 DATA SOURCES AND FIXED CAPITAL INVESTMENT BASES

The process data for the H-Coal liquefaction process and its yields were taken from a report prepared by HRI for U.S. DOE under Contract No. EX-76-C-01-1544 entitled "H-Coal Integrated Pilot Plant", Volumes I and II, November 1977.¹⁰ Minor adjustments to the yields were made to account for the use of Illinois No. 6 coal with a composition slightly different from that used in the referenced HRI report.

The H-Coal process catalytically converts coal to oil. At high severity conditions a liquid product called syncrude is produced. At low severity conditions, a low-sulfur, liquid fuel referred to as fuel oil is the product. This flexibility is achieved by varying the reactor space velocity which results in change of the hydrogen consumption per ton of coal. The yield of gas is reduced significantly when fuel oil is the primary objective rather than synthetic crude. However, the conversion of coal to liquids remains high even in the low severity fuel oil cases. The referenced HRI report presents data on yields and operating conditions from their PDU runs for both syncrude and fuel oil mode of operation. Anti-

cipated yield and run conditions for intermediate severities of operation are also reported.

Yields for this report are based on the projected pilot plant yields - intermediate mode, for Illinois No. 6 coal.¹⁰ The severity of operation for this intermediate mode is similar to the syncrude mode of operation and allows atmospheric distillation of the reactor effluent prior to the filtration step.

Basic data and yields for the gasification units originated from work done for OCR under Contract No. 14-32-0001-1207.⁹ Independent designs were developed by Parsons for the specific conditions of temperature, pressure, and gas composition used.

The bases for the constructed costs of the units shown in block flow diagram R-02-FS-1 were taken from the following:

1. A report prepared by The American Oil Company (AMOCO) for OCR under contract No. 14-01-0001-1188, entitled "Evaluation of Project H-Coal", December 1967.¹¹
2. A report prepared by Fluor Engineering and Construction, Los Angeles, California, in 1976 under Contract No. E(49-18)-2002 for ERDA entitled "H-Coal Commercial Evaluation-Conceptual Design and Economic Analysis for 25,000 TPD H-Coal Liquefaction Plant."¹²
3. A report prepared by Parsons for ERDA Under Contract No. E(49-18)-1775 published in March 1977, entitled "Oil/Gas Complex-Conceptual Design/Economic Analysis - Oil and SNG Production", R&D Report No. 114, Interim Report No. 4.⁶
4. A report prepared by Parsons for ERDA under Contract No. E(49-18)-1775 published in January 1977, entitled "Fischer-Tropsch Complex-Conceptual Design/Economic Analysis - Oil

and SNG Production," R&D Report No. 114, Interim Report No. 3.⁵

5. A report prepared by Parsons for OCR Under Contract No. 14-32-0001-1234 published in 1973 entitled "Demonstration Plant - Clean Boiler Fuels from Coal-Preliminary Design/Capital Cost Estimate", R&D Report No. 82, Interim Report No. 1.¹³
6. A report prepared by Parsons for DOE, under Contract No. EX-76-C-01-1775 published in December 1977 entitled "Project POGO-Coal Refinery Complex - Conceptual Design/Economic Analysis -Power-Oil-Gas-Other Products", R&D Report No. 114, Interim Report No. 6.²
7. Parsons in-house information.

5.2.3 PROCESS DESCRIPTION

The H-Coal process is shown schematically in block flow diagram R-02-FS-1. The heart of the process is the Dissolving and Separation unit which is a catalytic hydroliquefaction process. Here, a slurry of pulverized coal in a process-produced heavy recycle solvent made up of vacuum bottoms, filtrate and atmospheric bottoms is pumped to about 3,000 psig, then mixed with hydrogen. The coal slurry and hydrogen mix is preheated in a fired heater and then contacted with commercial catalyst¹⁰ in ebullating-bed reactors, which are operated at about 800°F and 3,000 psig. The catalyst is kept in suspension by the upflow of reactants and an internal liquid recycle. Make-up catalyst is continually added to the reactor and an equivalent amount of spent catalyst is continually withdrawn.

The reactor effluent slurry is sent to the atmospheric distillation unit to:

1. Separate the C₄ and lighter gases from the C₅+ liquids.

2. Separate the C_5+ liquids into:

- A. Naphtha (C_5 to 400° boiling range). Used as a feedstock to the naphtha hydrogenation unit.
- B. Light distillate (400 to 500°F boiling range). Used as a constituent of fuel oil products.
- C. Heavy distillate (500° - 650°F boiling range). Used both as wash oil in the filters and as a constituent of the fuel oil products.
- D. Atmospheric bottoms ($650^\circ\text{F}+$ slurry). Used both as a recycle slurring oil and vacuum filters feedstock.

The atmospheric bottoms are filtered to separate a filter cake, which contains unreacted coal and ash, from the liquefied coal products. The filtrate ($650^\circ\text{F}+$) is split into recycle slurring oil, filtrate product and vacuum distillation feedstock. The vacuum distillation unit produces:
(1) vacuum bottoms ($975^\circ\text{F}+$) which is used as a recycle slurring oil and
(2) vacuum distillate (650° - 975°F boiling range).

The Vacuum distillate is combined with the filtrate product ($650^\circ\text{F}+$) and part of the 400° - 650°F distillate. The resulting residual fuel oil ($400^\circ\text{F}+$) is projected to contain 0.3 Wt% sulfur and meet the commercially acceptable viscosity requirement of 180 SSF at 122°F . The remaining blend of light distillate (400° to 500°F) and heavy distillate product (500° to 650°F) meets the specifications of ASTM No. 4 fuel oil. The C_5 - 400°F naphtha is hydrotreated and catalytically reformed to produce high octane reformat.

The dissolving reactor product and recycle gases are combined with the atmospheric distillation gas and the naphtha hydrotreating of gas and sent to the high pressure acid gas removal unit.

After removal of H_2S and CO_2 the sweet gas is sent to the hydrogen recovery unit for separation and the acid gas is fed to the sulfur plant. The recovered hydrogen is recycled to the dissolver. The methane rich gas is sent to the SNG methanator for methanation of residual CO and inclusion in the SNG product. The C_2+ fraction is fed to the LPG fractionation unit where C_3 and C_4 LPGs plus light hydrocarbons are separated, the light ends are added to the SNG while the C_3 and C_4 LPGs are sent to storage.

The reactor product undissolved coal and ash, which is separated as filter cake from the reactor liquids, is combined with coal and gasified in a two-stage entrainment slagging gasifier. A low-Btu fuel gas is produced; after H_2S removal, this is used for plant fuel gas.

The hydrogen required for the process is generated by producing syngas at 1,000 psig in a two-stage entrained slagging gasifier. The syngas is shifted, the acid gases are removed, and a hydrogen rich gas is cryogenically separated. Residual CO in this stream is methanated to produce a high purity hydrogen make-up stream.

Syngas and methane-rich gas are combined with the C_2 and lighter gases produced in the hydrogen recovery unit and fed to a methanation unit to produce a high-Btu pipeline quality SNG.

The plant is a self-contained facility producing all of its own power, steam, and fuel gas; the only import utility item required is raw water.

5.2.4 ASSESSMENT: PROCESSING OTHER COALS

No information is available, at the time of this writing, on the yields from H-Coal when processing Eastern Coal Province (Pittsburgh) coal. Some information concerning its dissolving characteristics (under non-catalytic conditions) can be obtained from the bench-scale SRC II runs conducted by P&M at their Meriam Laboratory. The results from these runs indicate that the Pittsburgh Seam coal has MAF conversion similar to the

Illinois No. 6 coal; in addition, under similar reaction conditions it results in a higher hydrogen consumption, higher distillate yield and a lower SRC (650°F+) yield. Although the dissolver of the H-Coal process operates at a higher pressure and under catalytic conditions, it is reasonable to expect a similar change of the H-Coal yields when Pittsburgh is used instead of Illinois No. 6.

The net effect of this change would be a decreased coal and coal-surplus hydrogen feed rate to the dissolver. This change is due to the higher heating value of the feed coal while maintaining the total product energy output of the plant at 600 MM Btu/day. The decreased coal-sourced hydrogen feed rate will be offset by a higher molecular hydrogen consumption.

The results of gasifier elemental balances performed for several coals (in-house information) indicate that the process gasifier for the Eastern Coal Province coal would require less coal and oxygen in order to produce the same quantity of hydrogen and carbon monoxide in the syngas as that produced in the Illinois No. 6 (Interior Coal Province) case. The above changes would result in a decrease of the capital cost since the coal preparation unit, the coal dissolving unit and the oxygen plant would be smaller. There should be little change in the thermal efficiency since the decreased coal feed rate would be offset by the higher HHV of the Eastern coal.

Yield data for H-Coal processing of the Rocky Mountain coal were taken from a report entitled "H-Coal Integrated Pilot Plant" Volume II, 1977, prepared by Hydrocarbon Research, Inc. for DOE under Contract No. EX-76-C-01-1544.¹⁰ The data indicates that for similar conditions in the dissolver the MAF conversion of the Rocky Mountain Coal is lower (less than 90%) when compared with the Interior coal (95-95%). The relatively high oxygen content results in higher water and CO₂ yields. In order to maintain the plant product HHV output of 600 MMM Btu/day, more coal and more hydrogen will have to be fed to the dissolver. The gasifier elemental analyses also indicate that the Rocky Mountain feed rate, as well as the

oxygen feed rate, would have to be higher in order to produce the same quantity of ($H_2 + CO$) in the syngas as produced in the Interior case. Another disadvantage when processing Rocky Mountain coal is the high run-of-mine water content, which would require extensive drying facilities. The changes listed above will result in an increased capital cost for the Rocky Mountain coal compared to the Interior coal, due to the greater capacities of the units handling the increased coal, hydrogen and filter cake rates. Another effect of H-Coal processing of Rocky Mountain coal would be a lower thermal efficiency when compared to the Interior case. Economically, these factors tend to be offset by the lower cost of the western coal.

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5.3 CSF BASED PROCESS

5.3.1 DEVELOPMENT STATUS

Extensive bench-scale testing and research work was conducted by Consolidation Coal Company prior to 1963. The Office of Coal Research and Consolidation Coal entered into a contract in September 1963 for development of the CSF process. A 20-TPD pilot plant was designed, utilizing Consol's data, and built.

The pilot plant was operated for approximately 40 months during the 1966-1970 period and then was taken out of service. During this operating period, a number of equipment and mechanical problems were defined.

5.3.2 DATA SOURCES AND FIXED CAPITAL INVESTMENT

The CSF plant configuration used is illustrated by the block flow diagram shown in drawing R-03-FS-1. The process data and yields for the CSF process were taken from the DOE report entitled "Conceptual Design for Advanced Coal Liquefaction Commercial Plant," prepared under Contract No. EX-76-C-01-2251.¹⁴ That design/economic evaluation was adjusted to show projected performance when feeding Illinois No. 6 coal as an alternate to the Pittsburgh No. 8 seam coal used in the reference design. The reactor yields for the extraction, extract hydrogenation and low temperatures carbonization sections were concluded to be similar for the two coals. The percent heteroatom (N, S, O) removal in the extraction and extract hydrogenation sections was judged to be the same.

The data and yields for the gasification units originated from work done for OCR by BCR under Contract No. 14-32-0001-1207⁹ and Contract No. 14-01-0001-324.¹⁵ Independent designs were developed by Parsons for the temperature and pressure conditions specified for the CSF case.

The constructed costs of the units shown in the block flow diagram R-03-FS-1 were based on information from the following:

1. A report prepared for DOE under Contract No. EX-76-C-01-2251 published in October 1977, entitled "Conceptual Design for Advanced Coal Liquefaction Commercial Plant".¹⁶
2. A report prepared by Parsons for ERDA under Contract No. E(49-18)-1775 published in March 1977, entitled "Oil/Gas Complex -Conceptual Design/Economic Analysis - Oil and SNG Production", R&D Report No. 114 - Interim Report No. 4.⁶
3. A report prepared by Parsons for DOE under Contract No. EX-76-C-01-1775, published in December 1977 entitled "Project POGO - Coal Refinery Complex Conceptual Design/Economic Analysis", R&D Report No. 114 - Interim Report No. 6.²
4. Parsons in-house information.

5.3.3 PROCESS DESCRIPTION

The heart of the CSF coal liquefaction process includes the following process steps:

- Extraction
- Solids separation
- Low temperature carbonization
- Solvent recovery
- Extract hydrogenation

Minus- $\frac{1}{4}$ -inch coal is slurried with recycle solvent and recycle donor solvent and the resultant mixture is pumped to the preheater. The mixture is heated and fed to the extraction reactor, which is a stirred reactor operated at about 700°F and 575 psig and serves to extract coal

liquids from the feed coal. The hydrogenated donor solvent serves as a hydrogen carrier. After extraction, the solids are removed from the coal liquids in the solids separation unit which consists of several parallel gravity settling tanks.

Extraction solids, contained in the settler underflow stream, are sent to the low temperature carbonization (LTC) unit which is operated at about 400 psig and 900°F. Heat is supplied by partial oxidation of the undissolved coal with oxygen. Solvent contained in the settler underflow and LTC produced tar is recovered after separation from the LTC gases. Tar and recovered solvent rejoin the settler overflow. The LTC gases are sent to the H₂S removal unit to be subsequently used for plant fuel gas and the LTC char is blended with extract hydrogenation vacuum residue in the process gasifier to produce the plant hydrogen and fuel gas requirements.

Extraction liquids, as settler overflow, are fractionated in the solvent recovery unit to produce a distillate which is sent to fuel oil distillation, recycle solvent which is returned to the extraction, and extract as bottoms which is forwarded to the extract hydrogenation unit.

In the extract hydrogenation unit, the extract is reacted with hydrogen in a catalytic ebullating bed reactor at about 3,000 psig and 800°F. The gas phase is separated and sent to the acid gas removal unit and the resulting sweet gas is sent to the fuel gas cryogenic separation plant for separation into final products — SNG, C₃ LPG and C₄ LPG. The liquid phase, including the solids, is fractionated in atmospheric and vacuum columns to produce naphtha, light distillate, heavy distillate and vacuum bottoms. The vacuum bottoms, which contains the solids, is sent to the process gasifier to produce syngas. The light and heavy distillate are combined with the light oil from solvent recovery and sent to fuel oil distillation to produce the No. 4 and heavy fuel oil products. The naphtha is further refined by hydrodesulfurization and catalytic reforming to provide a high octane gasoline.

The hydrogen and fuel gas required for the process are produced by generating syngas from the gasification of LTC char and extract hydrogenation

tion vacuum residue. The fraction of syngas used for hydrogen production is shifted, the acid gases removed, and cryogenically separated into a hydrogen-rich stream plus feed streams to the SNG and LPG purification units.

The required fuel gas is produced by processing a portion of the process gasifier product to remove hydrogen sulfide. This fuel gas is then used to produce the steam and electrical power required to operate the complex.

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

5.3.4 ASSESSMENT OF PROCESSING OTHER COALS

The projected performance of the CSF process when feeding Appalachian and Rocky Mountain coals is summarized in Table 5-2. The data source for these projections are bench-scale and pilot plant experimental programs. The coal compositions are presented in this table to indicate the differences in compositions from those similar areas, listed in Table 3-3, used in this study. Although they differ, the coals are from the same geographical area and, assuming similar liquefaction tendencies, are useful for comparison.

As indicated in the table, the MAF conversions are very similar for both the extraction and the extract hydrogenation steps. The MAF conversion for all three coals is defined as:

$$\frac{(\text{MAF feed} - \text{the unconverted residue})}{\text{MAF feed}} \times 100.$$

However, the hydrogen consumption is significantly higher for the Montana coal, 4.6 % versus 4.0% and 3.8% for Pittsburgh No. 8 and Illinois No. 6 coals, respectively. Most of this can be attributed to the high oxygen content of the Montana coal which is typical of Western sub-bituminous coals.

A comparison of the yields indicated the Pittsburgh coal to have the higher liquid yields and lowest total gas yield. However, all of the coals have acceptable oil yields to be candidates for a commercial plant.

As previously stated for the SRC II case, the net effect of processing Rocky Mountain coal compared to the Pittsburgh No. 8 and Illinois No. 6 coal would be a lowering of thermal efficiency due to:

1. The higher drying load because of high moisture content in Rocky Mountain coal.
2. Additional hydrogen required to convert the oxygen to water due to the relatively high oxygen content.

This would result in a higher capital investment for the same quantity of fuel products in a commercially sized plant for the Rocky Mountain coal. However, the higher investment would tend to be compensated for by the significantly lower cost of Rocky Mountain coal compared to that for Eastern and Interior coals.

Table 5-2 - CSF Process Yields For Several Coals

Item	Eastern Coal Province ^a (Pittsburgh No. 8 Seam Ireland Mine)	Interior Coal Province ^b (Illinois No. 6)	Rocky Mountain Coal Province ^b (Montana Sub- bituminous)
Coal Composition (Dry Wt %)			
Carbon	69.0	64.8	69.8
Hydrogen	4.9	4.4	4.5
Nitrogen	1.3	1.2	1.0
Sulfur	4.3	5.8	0.9
Oxygen	7.3	8.3	14.4
Ash	<u>13.2</u>	<u>15.5</u>	<u>9.4</u>
	100.0	100.0	100.0
Yields (Wt % Dry Coal)			
H ₂ O	5.7	7.4	11.3
C ₁ -C ₄	9.1	7.7	9.5
Other Gases	7.1	12.7	10.9
Light Distillate (Naphtha)	6.2	8.7	9.2
Heavy Distillate	30.0	24.3	25.8
Hydro Extraction Residue (750°F+)	10.9	8.8	9.4
LTC Char	34.5	36.8	29.8
Dry Wt % H ₂ Consumption	3.5	3.2	4.2
MAF Wt % H ₂ Consumption	4.0	3.8	4.6
MAF Conversions			
Extraction (Wt % of MAF Coal)	70	68	69
Extract Hydroconversion (Wt % of MAF Feed)	81	83	82
Total Gas Yield (Wt % Dry Coal)	16.8	20.4	20.4
Total Distillate Oil Yield (Wt % Dry Coal)	36.2	32.0	35.0
Total Distillate Oil + Residue Yield (Wt % Dry Coal)	47.1	41.8	44.4
Residual Fuel Oil Yield (Heavy Distillate + Residue) (Wt % Dry Coal)	40.9	33.1	35.2

^aFoster Wheeler Report¹⁷ for OCR, R&D Report No. 70, 1972^bConsolidation Coal Co. Report for OCR, R&D Report No. 39, Vol I, 1969¹⁸

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5.4 FISCHER-TROPSCH BASED PROCESS

5.4.1 DEVELOPMENT STATUS

Development of Fischer-Tropsch technology by Drs. Fischer and Tropsch in Germany during the 1920s led to construction of the first commercial plant in 1936. In 1939, nine plants existed in Germany with a total rated capacity of about 750,000 mtpy of liquids.

SASOL I in the Republic of South Africa is a Fischer-Tropsch plant that is reported to have produced and processed in excess of 8.5 million standard cubic meters per day of syngas and has been in operation since 1955. A program is now under way to construct SASOL II which will be on the order of 4 to 5 times larger and was scheduled to begin operation in 1980.

Fischer-Tropsch chemistry is long established and proven. There is an incentive to develop plant designs with increased efficiency and improved economics. The Pittsburgh Energy Research Center (PERC) of DOE, the U. S. Bureau of Mines, and its predecessor R&D Activity Under has investigated the use of flame sprayed catalysts for use in Fischer-Tropsch and methanation conversions over a 15-year period. A conceptual design has indicated that this system has potential advantages.⁵ PERC is now initiating a small scale pilot plant test program to further investigate this method of reactor design and catalysis to determine if the potential advantages can be achieved.

5.4.2 DATA SOURCES AND FIXED CAPITAL INVESTMENT BASES

The data for the Fischer-Tropsch coal conversion process and its yields were taken from a report prepared by Parsons for ERDA.⁵ This report also supplied the bases for the fixed capital investment estimate. The coal conversion units shown in the block flow diagram also came from this report.

5.4.3 PROCESS DESCRIPTION

The Fischer-Tropsch (F-T) process is shown schematically in the block flow diagram designated as Drawing R-04-FS-1.

Here the prepared coal is gasified in two-stage, entrained, slagging-type gasifiers at about 470 psig. The product gas, consisting primarily of a mixture of carbon monoxide and hydrogen (syngas), is cleaned to remove particulates by scrubbers and electrostatic precipitators. Part of the syngas is then sent to an isothermal, sour shift conversion unit to adjust the H_2 to CO ratio to that required for the F-T synthesis reactor. The clean, shifted syngas is then treated in a selective acid gas removal unit to remove CO_2 and H_2S . The syngas sulfur content is then reduced to 0.05 ppm by zinc oxide/cobalt-molybdenum guard reactors before entering the synthesis reactors. The F-T reactors are extended surface (finned-tube) heat exchangers with a flame-sprayed iron oxide catalyst deposited on the outside surface. The exothermic reaction is controlled by the generation of high-pressure steam in the tubes; the ratio of cold recycle gas to fresh feed syngas is 1.5:1. The reactor operates at about 600°F and 400 psig.

Methane rich gases (C_2 and lighter), containing some unreacted CO and H_2 , are separated from the F-T liquids and sent to the methanation unit to produce SNG.

The liquids produced in the F-T synthesis step, which are essentially sulfur free, are separated in the liquid product recovery unit into C_3/C_4 , C_5 -400°F and 400°F-plus fractions.

The 400°F-plus liquid, which is highly paraffinic, is sent to the catalytic cracking unit to reduce the pour point. Heavy cycle oil, after fractionation, is sold as low-sulfur residual fuel oil; light cycle oil, after hydrotreating to saturate the liquids in order to reduce the tendency to form gum, is sold as low sulfur No. 2 fuel oil; light and heavy catcracked gasoline are sent to gasoline blending. The propylene and butylene produced by the catalytic cracking unit are combined with the propane and

butane stream from F-T synthesis; this stream is debutanized and then catalytically polymerized. Polymer gasoline is sent to gasoline blending. The propane and butane from catalytic polymerization are fractionated to produce propane for use in SNG and butane for LPG and gasoline blending.

The C_5 - 400°F naphtha, which is paraffinic and has a low octane number, is hydrotreated and catalytically reformed to produce a high octane C_5 -plus reformat and sent to gasoline blending.

The F-T synthesis unit produces a condensate stream which contains oxygenates, such as alcohols, aldehydes, ketones, and organic acids, plus water. The alcohol is separated in the chemical recovery unit from the organic acid and water stream.

The plant is a self-contained facility producing all of its own power, steam, and fuel gas. High pressure steam, which is generated in the process gasification unit, is used for plant steam requirements and to generate electricity for the plant power requirements. Significant quantities of high pressure steam are also produced in the shift conversion, F-T synthesis, and methanation units.

5.4.4 ASSESSMENT OF PROCESSING OTHER COALS

Table 5-3 compares gasification yields from several coals;² these in turn largely determine the impact on the F-T process.

The data indicate that for a constant quantity of syngas (H_2+CO) feed to the F-T synthesis step, the Eastern Coal Province coal and oxygen requirements would be about 9% and 5% less, respectively, than the Interior Coal Province coal. Therefore, the capital cost should decrease slightly since 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.

The data indicate, however, that the Rocky Mountain Coal Province coal and oxygen feed rates would each be about 10% greater than

for the Interior case to produce the same quantity of syngas. On this basis, the capital cost would increase slightly and the thermal efficiency would remain essentially unchanged since the lower HHV of the Rocky Mountain coal would offset the increased coal feed rate.

However, the Rocky Mountain Coal Province run-of-mine-coal has a water content of approximately 25% and an ash content of only about 6%. Therefore, the coal preparation facilities would be deleted since the ash content is sufficiently low; however, extensive drying facilities would be required to reduce the water content to 2.7%. Thus, the net change in processing Rocky Mountain coal would be a decrease in thermal efficiency and an increase in fixed capital cost. The higher capital cost is expected to be offset by the significantly lower coal cost than the Eastern and Interior coals.

Table 5-3 - Gasifier Yields For Several Coals

Item	Interior Coal Province	Eastern Coal Province	Rocky Mountain Coal Province
Coal Proximate Analysis ² (Wt %)			
Moisture	2.7	2.7	2.7
Ash	8.0	8.0	8.0
Volatile Matter	38.1	31.8	44.2
Fixed Carbon	51.2	57.5	45.1
	<u>100.0</u>	<u>100.0</u>	<u>100.0</u>
HHV, Btu/lb	12,425	13,590	11,211
Coal Ultimate Analysis ² (Wt %)			
Carbon	70.1	75.7	66.3
Hydrogen	4.8	4.8	4.7
Nitrogen	1.4	1.6	1.0
Sulfur	3.9	1.6	0.8
Oxygen	9.1	5.6	16.5
Moisture	2.7	2.7	2.7
Ash	8.0	8.0	8.0
	<u>100.0</u>	<u>100.0</u>	<u>100.0</u>
Gasifier Gas, Volume % (Dry) ^a			
Hydrogen, H ₂	40.8	41.6	39.9
Nitrogen, N ₂	0.9	0.9	0.8
Carbon Monoxide, CO	44.9	46.2	43.6
Carbon Dioxide, CO ₂	11.6	10.3	14.8
Hydrogen Sulfide, H ₂ S	1.2	0.4	0.3
Carbon Oxysulfide, COS	TR	TR	TR
Methane, CH ₄	0.6	0.6	0.6
	<u>100.0</u>	<u>100.0</u>	<u>100.0</u>
Relative Rates for Constant (H ₂ +CO) Production			
Coal Feed	1	0.91	1.13
Oxygen Feed	1	0.95	1.09
^a Calculated			

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	TONS/DAY	BPSD (MMSCFD)	MMMBTU/D (HHV)
FEED COAL	36,340	-	910
SNG (972 BTU/SCF)	7,275	52,665* (325)	316
BUTANES	1,185	11,750	51
GASOLINE	4,645	37,535	197
OXYGENATES	640	3,760	13
NO. 2 FUEL OIL	410	2,725	16
HEAVY FUEL OIL (0.0 WT% SULFUR)	170	1,075	7
SULFUR	1,230	-	-
TOTAL	15,555	109,510	600

*BARRELS FOE

THERMAL EFFICIENCY: $\frac{600}{910} \times 100 = 66\%$

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0			ISSUE FOR REPORT	SRD	FLC				
COAL LIQUEFACTION PROCESS SURVEY BLOCK FLOW DIAGRAM FISCHER-TROPSCH (FSC) PROCESS									
THE RALPH M. PARSONS COMPANY PASADENA, CALIFORNIA			JOB NO. 5658-1	DWG. NO. R-04-FS-1		REV. 0			

SECTION 6
ENVIRONMENTAL

SECTION 6

ENVIRONMENTAL

6.1 GENERAL

This section defines the environmental procedures and equipment included in the designs of the five high potential processes selected, and compares projected effluents with applicable and related environmental standards.

Two major environmental aspects were considered in detail: air emissions and water treatment and supply. These aspects are discussed at first independently for each process; they are then compared with each other and with applicable standards to determine the relative environmental acceptability of each high potential process. Other aspects, such as generation and disposal of solid wastes, biohazards of the intermediates and products, and environmental acceptability of the fuels produced are also briefly reviewed.

The same data base utilized for the "Description of the High Potential Processes" (Section 5) was used to develop quantitative projections for air emissions and water consumption and discharge. All processes were assumed to use an Illinois No. 6 coal (3.4% sulfur, 7.1% ash) as raw material.

6.2 AIR EMISSIONS

The major air pollution abatement effort is aimed at removing particulates and sulfur dioxide from the gases generated during the coal conversion process to make the fuels produced environmentally acceptable.

The air pollution abatement procedures followed for each process are outlined in block flow diagrams, which also show projections for the types and amounts of the streams vented to the air from specific units. Effluents from various units are often combined into a main plant stack; amounts and

concentrations emitted to the air from specific stacks for each process are detailed in a series of tables. Applicable standards and compliance of the individual gaseous emissions are discussed in the following sections.

6.2.1 SRC II PLANT

The generation and control of air emissions for the SRC II Plant are outlined in Figure 6-1, which also shows the nature and amount of all streams vented to the air. The effluent gases are shown vented separately to the air to identify the contribution of specific process units. However, the gaseous effluents from the selective acid gas removal unit and the tail gas treatment unit are combined in a single stack before venting to the air.

The coal drying unit is the only major source of particulate emission. A cyclone followed by a baghouse remove the majority of the particulates from the vent stream.

Particulates present in the gas stream generated by the process gasifier are removed by a cyclone followed by a dust filter. The solids collected (char and ash) are returned to the gasifier. After shift conversion the acid gases (mainly hydrogen sulfide, hydrogen cyanide, carbon oxysulfide and carbon dioxide) present in the process gasifier stream are separated from the other main syngas components (hydrogen, carbon monoxide, and methane) in a selective acid gas removal unit. On regeneration of the solvent, carbon dioxide with traces of carbon monoxide and hydrogen sulfide is separated from the other acid gases and vented to the air; hydrogen sulfide and other contaminants are conveyed to the sulfur recovery plant.

Acid gases originating from the dissolver and consisting mainly of hydrogen sulfide are removed by amine scrubbing in an acid gas removal unit. On regeneration of the amine, the hydrogen sulfide liberated is conveyed to the sulfur recovery plant.

The sulfur recovery plant receives the two hydrogen sulfide streams mentioned above, plus a similar stream from the sour water stripping

unit, and converts approximately 95% of the sulfur present to elemental sulfur; other contaminants are thermally oxidized to inoffensive compounds. The remainder of the sulfur is converted in a tail gas treatment unit, where sulfur species are reduced to hydrogen sulfide, absorbed by an alkaline redox solution, and oxidized to elemental sulfur. The final effluent, containing carbon dioxide plus traces of carbon oxysulfide, hydrogen sulfide, and carbon monoxide, is vented to the air.

The lower-Btu gas stream produced by the fuel gasifier is freed of particulates by a cyclone followed by an electrostatic precipitator; the solids collected (ash and char) are returned to the gasifier. Practically all remaining traces are removed by wet contact in the sulfur removal unit downstream. The hydrogen sulfide present is absorbed and converted to elemental sulfur in this unit using a redox process analogous to the one employed in the tail gas treatment unit mentioned above. The cleaned fuel gas is utilized for steam generation and process heating; on combustion of the fuel gas, carbon dioxide, sulfur dioxide, and nitrogen oxides are emitted to the air.

Amounts and concentrations of the combined effluents emitted to the air are detailed in Table 6-1, while the sulfur balance for the plant is presented in Table 6-2. Compliance with applicable standards is discussed in Section 6.2.6.

6.2.2 H-COAL PLANT

The dissolver in the H-Coal Plant employs a catalytic process, while the SRC II employs a psuedo-catalytic technology. Other key units employed in the H-Coal and in the SRC II plants serve similar purposes. The generation and control of air emissions for the H-Coal Plant are outlined in Figure 6-2; amounts and concentrations of the combined effluents emitted to the air are detailed in Table 6-3, while the sulfur balance is presented in Table 6-4. The different dissolving process is reflected in differing ratios of individual pollutants when composed with the SRC II plant.

6.2.3 CSF PLANT

The generation and control of air emissions for the CSF Plant are outlined in Figure 6-3, while it also shows the nature and amount of all streams vented to the air. The effluent gases are shown vented separately to the air to identify the contribution of specific process units. However, the gaseous effluents from the acid gas removal unit and the tail gas treatment unit are combined in a single stack before venting to the air.

The coal preparation unit is the only major source³ of particulate emissions. A cyclone followed by a baghouse remove most of the particulates from the vent stream.

Particulates present in the gas stream generated by the process gasifier are removed by a cyclone followed by an electrostatic precipitator. The solids collected (ash and char) are returned to the gasifier. The gas stream is then split into two fractions. One fraction undergoes shift conversion, then 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 vented to the air; hydrogen sulfide and other contaminants are conveyed to the sulfur plant.

The other fraction is conveyed to a sulfur removal unit where the hydrogen sulfide present is absorbed by an alkaline redox solution and oxidized to elemental sulfur. The cleaned lower-Btu gas produced is used as plant fuel, in process heaters and steam boilers; on combustion of the fuel gas, carbon dioxide, sulfur dioxide, and nitrogen oxides are emitted to the air.

Gases produced on extraction and low temperature carbonization are treated similarly in a hydrogen sulfide removal unit. The desulfurized gas is also used as plant fuel.

The sulfur plant receives the hydrogen sulfide stream from the acid gas removal unit and converts approximately 95% of the sulfur present to elemental sulfur; other contaminants are thermally oxidized to inoffensive compounds. The remainder of the sulfur is converted in a tail gas treatment unit, where sulfur species are reduced to hydrogen sulfide, absorbed by an alkaline redox solution, and oxidized to elemental sulfur. The final effluent, containing carbon dioxide plus traces of carbon oxysulfide, hydrogen sulfide, and carbon monoxide, is vented to the air.

Amounts and concentrations of the combined effluents emitted to the air are detailed in Table 6-5, while the sulfur balance of the plant is presented in Table 6-6. Compliance with applicable standards is discussed in Section 6.2.6.

6.2.4 FISCHER-TROPSCH PLANT

The generation and control of air emissions for the Fischer-Tropsch (F-T) Plant are outlined in Figure 6-4, which also shows the nature and amount of all streams vented to the air. The effluents are shown vented separately to the air to identify the contribution of specific process units. However, all gaseous effluents are combined in a single plant stack before venting to the air.

The coal drying unit is the only major source of particulate emissions. A baghouse removes most of the particulates from the vent stream.

In-plant energy requirements for process heaters and steam boilers are satisfied by steam produced by the F-T process. Therefore, no emissions from combustion of fuel gas are generated. There is only one gasifier, the process gasifier; the gas stream produced is freed of particulates by an abatement system consisting of a cyclone followed by a hot electrostatic precipitator; the solids collected (ash and char) are returned to the gasifier. Practically all of the remaining particulates are removed downstream by wet scrubbing during condensate removal and passage through cold electrostatic precipitators. The gas stream is then desulfurized by

an acid gas removal unit; the hydrogen sulfide separated is conveyed to a sulfur recovery and then tail gas treatment plant; these units operate similarly to the units described above for the SRC II and CSF plants. The last traces of sulfur present in the cleaned gas are removed by a zinc oxide guard bed, to avoid deactivating the F-T catalyst. Passage through the F-T catalytic reactors converts the synthesis gas to sulfur and nitrogen free liquid hydrocarbons and alcohols. Some CO₂ is also produced, removed by a CO₂ removal unit and vented to the air together with minor amounts of hydrocarbons and carbon monoxide.

Amounts and concentrations of the combined effluents emitted to the air are detailed in Table 6-7, while the sulfur balance for the plant is presented in Table 6-8. Compliance with applicable standards is discussed in Section 6.2.5.

6.2.5 COMPLIANCE WITH APPLICABLE SOURCE EMISSION STANDARDS

Source emission standards for coal conversion plants have not been issued by the Federal Government. However, emission standards for Coal Thermal Dryers, present in some coal conversion processes have been issued, and source emission standards for Petroleum Refinery Sulfur Recovery Plants, of a type similar to the ones employed in coal conversion technology, have been proposed.

Of the states, only New Mexico has issued specific regulations covering coal gasification plants. These regulations can be considered for illustrative purposes only because New Mexico is not included in the three coal source areas, namely the Appalachian Region of the Eastern Coal Province, the Eastern Region of the Interior Coal Province, and the Powder River Region of the Rocky Mountain Coal Province, selected in Section 3 as typical coal sources for this study. Three states representative of these regions are West Virginia, Illinois, and Wyoming; these states have issued standards for petrochemical and chemical manufacturing processes, which are related to coal conversion technology.

Applicable Federal, Illinois, New Mexico, West Virginia, and Wyoming source emission standards are listed in Table 6-9. The projected air emissions of the five high potential coal liquefaction processes are listed in Table 6-10; concentrations have been calculated on a dry basis to conform with the requirements of the standards. Two of the processes, CSF and Fischer-Tropsch are projected to meet all emission standards. The two other processes are projected to exceed some of the standards. The SRC II Plant is projected to exceed the allowable hydrogen sulfide emissions limits and the H-Coal Plant is projected to exceed the allowable nitrogen oxide emissions limits. It is believed that application of additional control technology presently available could bring these two processes into compliance with applicable standards. For example, use of countercurrent two-stage solvent and gas flows in the acid gas removal unit could reduce hydrogen sulfide emissions for the SRC II plant, while combustion modification could decrease the nitrogen oxides generated by the H-Coal process boilers.

6.3 WATER TREATMENT AND SUPPLY

Water treatment and supply are major environmental concerns for the five high potential coal liquefaction processes considered. It is assumed that for all processes the water is supplied by a nearby river; this water is pumped and used for cooling water makeup and boiler feedwater makeup. The progressive treatment and disposition of the aqueous effluent streams is considered for each process; the procedures involved are outlined in block flow diagrams, which also show water flow rates for the major streams. Water consumption is discussed in a separate section.

6.3.1 SRC II PLANT

The generation and treatment of aqueous contaminants for the SRC II Plant is outlined in Figure 6-5. Wastewater sources are listed on the left hand side of the figure, with the degree of pollution of the wastewater streams decreasing from top to bottom. Treatment and disposition of the streams is also shown; approximate flow rates are reported. The water supply system is outlined on the right hand side. 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 most highly contaminated stream is the sour water generated by several process units, such as the gasifiers, the dissolver, and the product distillation and hydrotreating units. This stream contains hydrogen sulfide, ammonium sulfide, phenols, cresols, thiocyanates, cyanides and solids (ash and char). Treatment consists of stream stripping to remove hydrogen sulfide and ammonia; the aqueous stream is then used as reagent water in the gasifiers, where organic contaminants undergo thermal destruction. Filters on return condensate streams remove the solids present.

Blowdown waters from the cooling tower and the boilers, containing small amounts of corrosion inhibitors (zinc salts and inorganic phosphates) are combined with soluble inorganic salts from the deionizer wastewater, neutralized, and treated with coagulants to precipitate suspended solids. The clean effluent water is then used for slag quench and dust control as required with the excess returned to the river.

6.3.2 H-COAL PLANT

The purification of raw river water supply and the generation and treatment of aqueous contaminants for the H-Coal Plant is outlined in Figure 6-6. Except for flow rates, the water supply and water treatment system is identical with the one discussed above for the SRC II Plant.

6.3.3 CSF PLANT

The purification of raw river water supply and the generation and treatment of aqueous contaminants for the CSF Plant is outlined in Figure 6-7. Except for flow rates, the water supply and water treatment system is identical with the one discussed above for the SRC II Plant.

6.3.4 FISCHER-TROPSCH PLANT

The generation and treatment of aqueous contaminants for the Fischer-Tropsch Plant is outlined in Figure 6-8. Wastewater sources are listed on the left hand side of the figure, with the degree of pollution of the wastewater streams decreasing from top to bottom. Treatment and disposition of the streams is also shown; approximate flow rates are reported. The water supply system is outlined on the right hand side. 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 most highly contaminated stream is the sour water generated by several process units, such as the gasifiers, the product distillation and hydrotreating units. This stream contains hydrogen sulfide, ammonium sulfide, phenols, thiocyanates, trace quantity of cyanides and solids (ash and char). Treatment consists of steam stripping to remove hydrogen sulfide and ammonia; the aqueous stream then undergoes wet oxidation with conversion of the organic materials to carbon dioxide, nitric oxide, and sulfur dioxide. These gases are conveyed to the gasifier, where nitric oxide is reduced to nitrogen and ammonia and sulfur dioxide is reduced to hydrogen sulfide. The purified aqueous stream is deionized and used for boiler feedwater makeup.

A second aqueous effluent stream containing large amounts of organics (alcohols, aldehydes, and organic acids) is produced on condensation of the Fischer-Tropsch synthesis product stream. This stream undergoes chemical treatment with caustic, then is combined with boiler water blowdown, and concentrated in a triple effect evaporator. The evaporator condensate is used for boiler feedwater makeup, while the residue is sprayed on the feed coal to the gasifier.

Blowdown water from the cooling tower, containing small amounts of corrosion inhibitors (zinc salts and inorganic phosphates) is combined with soluble inorganic salts from the deionizer wastewater, neutralized,

and treated with coagulants to precipitate suspended solids. The clean effluent water is then used for slag quench and dust control as required, with the excess returned to the river.

6.3.5 APPLICABLE AQUEOUS EFFLUENT STANDARDS

No aqueous effluent standards specifically addressed to coal-conversion plants have been issued by the Federal Government, or by state legislatures. Standards that are somewhat related to coal conversion processes are the Federal standards issued for petroleum refining. Maximum concentrations which were the base for Petroleum Refining New Source Performance Standards are reported in Table 6-11; discharges from the plants are expected to meet these standards. Actual information on the composition of the effluent streams is not available.

Many states have issued standards applicable to all sources discharging to the natural waters of the state. For illustration purposes, standards issued by the State of Illinois are reported in Table 6-12. Aqueous effluents from the plants would also have to meet these or similar state standards.

6.3.6 WATER CONSUMPTION

Water requirements for the four high potential processes considered are listed and compared graphically in Figure 6-9. The requirements are lowest for the SRC II, H-Coal and the Fischer-Tropsch processes and nearly 50% higher than their average for the CSF process. As shown in the figure, most of the water requirements are for cooling water, rather than for boiler and process water. A decrease in cooling water requirements could be obtained using wet-dry cooling towers in place of the wet cooling towers planned for the base case location.

6.4 SOLID WASTES

Large amounts of ash and slag (approximately 200 TPD for each plant) are expected to be generated. The available disposal procedures for these

waste products are (1) return to the mine for burial with the mine spoils or (2) discard in a suitable land fill. Contact of these materials with aquifers should be avoided, due to the possibility of leaching of metals.

Wastewater treatment procedures produce treatment sludges which may be contaminated with metals (e.g. zinc used as corrosion inhibitor). These sludges could also be disposed of at a mine site or landfill, as long as the possibility of ground or surface water contamination by leaching, is excluded.

6.5 BIOHAZARDS

Of particular interest in coal conversion projects is the possible formation of carcinogenic compounds on hydrogenation and pyrolysis of coal. These compounds are polynuclear aromatic hydrocarbons and heterocyclics usually found in coal tar.

Under the operating conditions of the entrained slagging type coal gasifiers used in the four processes considered, nil coal oil and coal tars are expected to be produced. The entire dissolver trains in the SRC II and H-Coal plants, and the solvent extraction train in the CSF Plant, however, handle coal-derived oils, and will therefore require strict application of industrial hygiene practices.

No dissolvers or solvent extraction units are included in the design of the Fischer-Tropsch plant. For this process, therefore, no coal-derived carcinogenic hazards are expected.

6.6 ENVIRONMENTAL ACCEPTABILITY OF THE FUEL PRODUCTS

Under authority of the Clean Air Act, the Environmental Protection Agency has issued regulations concerning the registration of commercial fuels;¹⁹ one of the requirements is testing for potential public health effects, including carcinogenic effects. This aspect has been discussed in the previous section.

The regulations also cover air emissions generated on combustion of fuels. The synthetic fuel oils produced by direct liquefaction (SRC II, H-Coal, and CSF processes) may contain 1 to 1.6% nitrogen; the amount of nitrogen oxides produced on combustion is therefore expected to be higher than for corresponding petroleum derived fuels (most common nitrogen content, 0.01 to 0.2%).

The same synthetic fuel oils also contain significant amounts of sulfur (up to the 0.5% value of SRC II fuel oil), which may prohibit their use in low air quality areas. For example, SRC II fuel oil and H-Coal residual fuel oil (0.3% sulfur) would not be acceptable in Southern California, where the maximum sulfur content allowed in fuel oil is 0.25%.

The fuels produced by indirect liquefaction (Fischer-Tropsch) on the other hand, contain nil sulfur and nitrogen.

Table 6-1 - Combined Effluents Emitted to the Air, SRC II Plant

Outlet	Gaseous Effluent	Amount	Concentration (ppm)
Main Stack	Carbon Dioxide	13,871 TPD	-
	Carbon Monoxide	6.2 TPD	565
	Carbon Oxysulfide (COS)	0.51 TPD	21.7
	Hydrogen Sulfide (H ₂ S)	321 lb/day	12.0
Process Heaters and Steam Boilers	Carbon Dioxide	14,480 TPD	-
	Sulfur dioxide	130 lb/day	0.6
	Nitrogen oxides (as NO ₂)	5 TPD	62.6
Coal Preparation Plant	Particulates	0.8 TPD	-

Table 6-2 - Sulfur Balance, SRC II Process

Item	TPD
Total Input from the Typical Feed Coal	1,041.0
Outputs:	
Recovered as Elemental Sulfur	988.2
As Reduced Sulfur Emissions (36% H ₂ S, 64% COS)	0.4
As Sulfur Dioxide Emissions from Combustion of Fuel Gas	0.03
In the Liquid Fuel Products	<u>52.4</u>
Total	1,041.0

Table 6-3 - Combined Effluents Emitted to the Air, H-Coal Plant

Outlet	Gaseous Effluent	Amount	Concentration (ppm)
Main Stack	Carbon Dioxide	18310 TPD	-
	Carbon Monoxide	1.7 TPD	107
	Carbon Oxysulfide (COS)	0.61 TPD	17.8
	Hydrogen Sulfide (H ₂ S)	205.9 lb/day	5.3
Process Heaters and Steam Boilers	Carbon Dioxide	12,701 TPD	-
	Sulfur Dioxide	117 lb/day	0.6
	Nitrogen Oxides (as NO ₂)	4.7 TPD	67
Coal Preparation Plant	Particulates	0.94 TPD	-

Table 6-4 - Sulfur Balance, H-Coal Process

Item	TPD
Total Input from the Typical Feed Coal	1,096.7
Outputs:	
Recovered as Elemental Sulfur	1,073.8
As Reduced Sulfur Emissions (23% H ₂ S, 77% COS)	0.4
As Sulfur Dioxide Emissions from Combustion of Fuel Gas	0.03
In the Liquid Fuel Products	<u>22.5</u>
Total	1,096.7

Table 6-5 - Combined Effluents Emitted to the Air, CSF Plant

Outlet	Gaseous Effluent	Amount	Concentration (ppm)
Main Stack	Carbon Dioxide	13753 TPD	-
	Carbon Monoxide	18.47 TPD	1774
	Carbon Oxysulfide (COS)	0.28 TPD	12.5
	Hydrogen Sulfide (H ₂ S)	204 lb/day	8.1
Process Heaters and Steam Boilers	Carbon Dioxide	16220 TPD	-
	Sulfur Dioxide	100 lb/day	0.6
	Nitrogen Oxides (as NO ₂)	6 TPD	96
Coal Preparation Plant	Particulates	0.95 TPD	-

Table 6-6 - Sulfur Balance, CSF Process

Item	TPD
Total Input from the Typical Feed Coal	1,214.3
Outputs:	
Recovered as Elemental Sulfur	1,201.1
As Reduced Sulfur Emissions (39% H ₂ S, 61% COS)	0.2
As Sulfur Dioxide Emissions from Combustion of Fuel Gas	0.2
In the Liquid Fuel Products	<u>13.0</u>
Total	1,214.3

Table 6-7 - Combined Effluents Emitted to the Air, Fischer-Tropsch Plant

Outlet	Gaseous Effluent	Amount	Concentration (ppm)
Plant Stack	Carbon Dioxide	51,660 TPD	-
	Carbon Monoxide	11.81 TPD	320
	Carbon Oxysulfide (COS)	1.7 TPD	21.5
	Hydrogen Sulfide (H ₂ S)	316 lb/day	5.5
	C ₂ -C ₆ Hydrocarbons (as CH ₄)	1.33 TPD	63
Coal Preparation Plant	Particulates	0.97 TPD	-

Table 6-8 - Sulfur Balance, Fischer-Tropsch Process

Item	TPD
Total Input from the Typical Feed Coal	1,235.6
Outputs:	
Recovered as Elemental Sulfur	1,234.6
As Reduced Sulfur Emissions (14% H ₂ S, 86% COS)	<u>1.0</u>
Total	1,235.6

Table 6-9 - Federal, Illinois, New Mexico, West Virginia,
and Wyoming Source Emission Standards

Pollutant	Federal Standards	Illinois Standards (Petrochemical Plant)	New Mexico Standards (Coal Gasification Plant)	West Virginia Standards	Wyoming Standards
Particulate Matter					
Volumetric Rate	0.031 gr/ft ³ ^a	-	0.03 gr/ft ³	0.07 gr/ft ³ ^a	-
Mass Rate (general)	-	E = 24.8 (p0.16) ^b	-	-	E = 17.31 (p0.16) ^b
Mass Rate (SRC II)	-	77.9 lb/hr	-	-	54.4 lb/hr
Mass Rate (H-Coal)	-	78.5 lb/hr	-	-	54.8 lb/hr
Mass Rate (CSF)	-	79.8 lb/hr	-	-	55.7 lb/hr
Mass Rate (Fischer-Tropsch)	-	80.0 lb/hr	-	-	55.9 lb/hr
Carbon Monoxide	-	200 ppm, 50% excess air	-	-	-
Organics (methane excluded)	-	100 ppm (CH ₄ equivalent)	-	-	-
Sulfur Dioxide	250 ppm ^c	1.2 lb/MM Btu	-	2000 ppm	2000 ppm
Nitrogen Oxides	-	0.7 lb/MM Btu	-	-	-
Hydrogen Sulfide					
Emissions	10 ppm ^c	-	10 ppm	-	-
Fuel Content	0.10 gr/ft ³ ^c	-	-	0.50 gr/ft ³	-
Total Reduced Sulfur (H ₂ S + COS + CS ₂)	300 ppm ^c	-	100 ppm	-	-
Hydrogen Cyanide	-	-	10 ppm	-	-
Hydrogen Chloride/Hydrochloric Acid	-	-	5 ppm	145 ppm	-
Ammonia	-	-	25 ppm	-	-
Gas Burning Process Boilers, Particulate Matter	-	-	0.03 lb/MM Btu, LHV	-	-
Gas Burning Process Boilers, Sulfur Dioxide	-	0.03 lb/MM Btu, HHV	0.16 lb/MM Btu, LHV	-	-
Gas Burning Process Boilers, Nitrogen Oxides	0.2 lb/MM Btu	0.2 lb/MM Btu	-	-	0.2 lb/MM Btu
Coal Burning Process Boilers, Particulate Matter	-	0.10 lb/MM Btu	-	-	0.10 lb/MM Btu
Coal Burning Process Boilers, Sulfur Dioxide	-	1.2 lb/MM Btu	-	1.6 lb/MM Btu	0.2 lb/MM Btu
Coal Burning Process Boilers, Nitrogen Oxides	0.7 lb/MM Btu	0.7 lb/MM Btu	-	-	0.7 lb/MM Btu
Total Sulfur	-	-	0.008 lb/MM Btu of feed	-	-
^a Standard for coal thermal dryers ^b E = emission rate, lb/hr; P = process input weight (ton/hour) ^c Proposed standard, petroleum refinery sulfur recovery plant (41 FR 43866, Oct. 4, 1976). ^d See particulate mass rate above					

Table 6-10 - Air Emissions, Coal Liquefaction Plants

Pollutant	SRC II	H-Coal	CSF	Fischer-Tropsch
Particulate matter:				
Volumetric Rate	0.03 gr/ft ³	0.03 gr/ft ³	0.03 gr/ft ³	0.03 gr/ft ³
Mass Rate	66.7 lb/hr	78.3 lb/hr	79.2 lb/hr	80.8 lb/hr
Carbon Monoxide	565 ppm ^a	107 ppm ^a	1774 ppm ^a	320 ppm ^a
Organics (methane excluded)	nil	nil	nil	63.1 ppm ^b
Sulfur Dioxide	nil	nil	nil	nil
Nitrogen Oxide	nil	nil	nil	nil
Hydrogen Sulfide:				
Emissions	<u>12.0 ppm</u>	5.3 ppm	8.1 ppm	3.5 ppm
Fuel Content	0.0004 gr/ft ³	0.0004 gr/ft ³	0.0004 gr/ft ³	nil
Total Reduced Sulfur (H ₂ S + COS + CS ₂)	33.7 ppm	23.1 ppm	23.1 ppm	20.1 ppm
Hydrogen Cyanide	nil	nil	nil	nil
Hydrogen Chloride/ Hydrochloric Acid	nil	nil	nil	nil
Ammonia	nil	nil	nil	nil
Gas Burning Process Boilers, Particulate Matter	nil	nil	nil	c
Gas Burning Process Boilers, Sulfur Dioxide, HHV	0.002 lb/MM Btu	0.005 lb/MM Btu	0.001 lb/MM Btu	c
Gas Burning Process Boilers, Sulfur Dioxide, LHV	0.002 lb/MM Btu	0.005 lb/MM Btu	0.001 lb/MM Btu	c
Gas Burning Process Boilers, Nitrogen Oxides	1.12 lb/MM Btu	<u>0.40 lb/MM Btu</u>	0.11 lb/MM Btu	c
Coal Burning Process Boilers, Particulate Matter	c	c	c	c
Coal Burning Process Boilers, Sulfur Dioxide	c	c	c	c
Coal Burning Process Boilers, Nitrogen Oxides	c	c	c	c
Total Sulfur	0.001 lb/MM Btu	0.001 lb/MM Btu	0.001 lb/MM Btu	0.002 lb/MM Btu
^a If the 50 excess air correction mentioned in the standard is applied, these values become: SRC II, 80 ppm, H-Coal, 20 ppm, CSF, 205 ppm Fischer-Tropsch, 126 ppm. ^b Methane equivalent ^c Not applicable because not included in process design. ^d <u>Underlined values</u> exceed source emission standards.				

Table 6-11 - New Source Performance Standards
for the Petroleum Refining Industry^a

Parameter	Maximum Concentration ^b (mg/l)
BOD-5	10
COD	60
Total Organic Carbon	21
Suspended Solids	6
Oil and Grease	3
Ammonia-N	10
Phenol	0.06
Sulfide	0.06
Chromium, Tertiary	0.16
Chromium, Hexavalent	0.003
^a Based on application of Best Available Demonstrated Technology (BADT). ^b Converted from the mass standards reported in: "Development Document for Effluent Guidelines and New Source Performance Standards for the Petroleum Refining Point Source Category." United States Environmental Protection Agency, Report EPA-440/1-74-014a, Washington, D.C., April 1974 (p. 145, 147, 176), Petrochemical Subcategory.	

Table 6-12 - Aqueous Effluent Standards, State of Illinois

Constituent	Maximum Concentration (mg/l)
Arsenic (total)	0.25
Barium (total)	2.0
BOD 5	10.0
Cadmium (total)	0.15
Chromium (total hexavalent)	0.5
Chromium (total trivalent)	1.0
Copper (total)	1.0
Cyanide	0.025
Fluoride (total)	15.0
Iron (total)	2.0
Iron (dissolved)	0.5
Lead (total)	0.1
Manganese (total)	1.0
Mercury (total)	0.0005
Nickel (total)	1.0
Oil (hexane solubles or equivalent)	15.0
pH	range 5-10
Phenols	0.3
Selenium (total)	1.0
Silver	0.1
Zinc (total)	1.0
Total Suspended Solids	12.0

Figure 6-1 - Block Flow Diagram
 Air Pollution Abatement, SRC II Plant

Figure 6-2 - Block Flow Diagram,
Air Pollution Abatement, H-Coal Plant

Figure 6-3 - Block Flow Diagram,
Air Pollution Abatement, CSF Plant

Figure 6-4 - Block Flow Diagram,
Air Pollution Abatement, Fischer-Tropsch Plant

Figure 6-5 - Block Flow Diagram,
Water Treatment and Supply, SRC II Plant

Figure 6-6 - Block Flow Diagram,
Water Treatment and Supply, H-Coal Plant

Figure 6-7 - Block Flow Diagram,
Water Treatment and Supply, CSF Plant

Figure 6-8 - Block Flow Diagram,
Water Treatment and Supply, Fischer-Tropsch Plant

Figure 6-9 - Comparison of Water Requirements for the Various Processes

SECTION 7
ECONOMICS

SECTION 7

ECONOMICS

This section summarizes the economic evaluation procedures used to analyze the high potential coal liquefaction technologies; it also presents the results of this evaluation. All economic factors are presented in first quarter 1978 dollars.

The economic comparison of four high potential processes includes individual estimates of projected fixed capital investment, total capital investment, operating costs, and profitability measures. Since the primary objective is development of relative economics, all cost and profitability comparisons are presented as ratios. The process with the lowest required capital investment, required product selling price (RPSP) or revenue was assigned a value of 1.0 and the equivalent value for all other processes for the particular economic factor under study was presented as a ratio to the reference process for that economic factor category. This procedure will become clearer when we begin to discuss the economic comparisons in detail.

For those readers who may want to convert the relative economic factors presented in this report section to the actual dollar values estimated in this study, the projected economics for SRC II are summarized in Table E-1 located in Appendix E.

The results of the Base Case analysis, consisting of the Eastern Region of the Interior Coal Province location, will be presented first. The impact of locating the plants in the Appalachian Region of the Eastern Coal Province and the Powder River Region of the Rocky Mountain Province will then be summarized. Finally, a comparison of projected economics for product supply to the four market areas - Delaware, Chicago, Houston, and San Francisco - will be presented. Sensitivities to key project parameters will also be summarized.

7.. PROCEDURES

The economic assessments are based on the process configurations previously described in Section 5 and 6 of this report, coupled with the procedures used to develop the comparative economic assessment as described briefly in the following paragraphs.

7.1.1 FIXED CAPITAL INVESTMENT

The fixed capital investment (FCI) source information for each process has been listed in Section 5. This includes experience generated by Parsons during the development of seven comprehensive conceptual designs plus the results of similar conceptual designs by other contractors. Parsons design projects required the development of preliminary specifications and cost data for over 6,000 major equipment items used in coal conversion plants. Using this equipment cost data, fixed capital estimates were developed for 80 units common to coal liquefaction plants such as coal preparation, acid gas treatment, sulfur removal, fractionation, filtration, and others. This information was used where appropriate to develop the FCIs for this study.

For analytical purposes, the coal conversion complexes were divided into the following four accounts/sections:

- Coal preparation, storage, grinding, and drying
- Process operations
- Utilities
- Facilities

In addition, 12.5% of the constructed cost total for these four accounts sections was added as an allowance for the costs of engineering, home office, fee, and sales tax.

The FCI for a given process plant used a combination of a number of types of estimating procedures as described below.

7.1.1.1 Coal Preparation, Storage, Grinding and Drying

The FCIs for these units were developed based on analysis of the results of similar units contained in the nine conceptual designs/economic evaluations generated by major contractors. Where required, the units were adjusted, scaled, and escalated.

7.1.1.2 Process Operations

The primary coal conversion units such as the coal hydroliquefaction, coal extraction, coal catalytic hydroliquefaction, gasification, and production of liquids from synthesis gas are specific to the separate processes. The FCIs were taken with judgement applied from major contractors' conceptual designs. Where required, the sizes of the units were scaled up or down. The costs were escalated to a first quarter 1978 basis. There are a large number of other types of units common to coal conversion plants; examples include acid gas removal units, sulfur plants, oxygen plants, filtration units, fractionation units, and product upgrading/refining units.

For the upgrading/refining units, FCIs were obtained by careful analysis and judgemental adjustment of prior inhouse and petroleum industry refining experience plus, where appropriate, use of experimental data generated while processing coal-derived liquids in refinery type operations.

An example of use of prior inhouse refining experience is catalytic reforming as a step to produce gasoline. Hydrotreating is an example of use of experimental data to define capacity and economics of a refinery type operations when processing coal-derived liquids. Here the experimental data developed in DOE/ERDA/OCR sponsored programs were used to establish design parameters such as space velocity, hydrogen consumption, and others. For the other important supporting process units such as acid

gas removal units, etc., the FCIs were developed by analysis and adjustment of prior related conceptual designs/economic evaluations.

7.1.1.3 Utilities

A utilities balance was developed for each candidate process. The FCIs for the utilities section was then developed from unit costs of prior conceptual designs/economic evaluations for related processes. Each case was analyzed prior to establishment of the FCI and suitable adjustments, and escalations made.

7.1.1.4 Offsites and General Facilities

Allowances for offsites and general facilities were developed based on the costs of similar facilities in related prior conceptual designs. The source costs were escalated to first quarter 1978 dollars.

Offsites and ancillaries include buildings, roads, utilities, communications, rolling stock, and related facilities required to efficiently service the plant operation and population.

7.1.2 TOTAL CAPITAL REQUIREMENTS

In addition to fixed capital investment, total capital requirements include initial charges of catalysts and chemicals, start-up costs, construction financing, and provision for working capital. Procedures for developing these costs are summarized below.

7.1.2.1 Initial Charges of Catalysts and Chemicals

The cost of the initial charge of catalyst and chemicals (C & C) was developed for each process. To develop the costs, each unit was analyzed and the quantities of initial C & Cs required were estimated and priced. The annual consumption of C & Cs was also estimated.

7.1.2.2 Start-up Costs

An allowance of 5% of FCI was used for start-up costs.

7.1.2.3 Construction Financing

The construction financing was calculated based on a five-year construction period, with the rate of funds drawdown as follows:

<u>Year</u>	<u>Percent Drawdown</u>	<u>Cumulative Percent Drawdown</u>
1	8	8
2	17	25
3	26	51
4	29	80
5	20	100

The cost of construction financing is based on 9% interest for funds drawn and a loan commitment fee of 0.75% for funds committed but not yet drawn.

7.1.2.4 Working Capital

Working capital consisted of the cost of:

- 15 days feed coal inventory
- 30 days inventory of liquid product
- Cost of spare parts inventory = 1% of FCI
- 30 days accounts receivable
- 30 days budget for current expenses

- 30 days credit for accounts payable

Working capital funds were borrowed for the project operating life.

7.1.3 OPERATING COSTS

Operating costs consist of the cost of purchased coal, materials and supplies, operating labor, maintenance, G & A overhead, property taxes and insurance, and miscellaneous. Methods of estimating these factors are described below.

7.1.3.1 Coal Cost

The coal conversion plants are conceived to be located close to mines. The analysis is based on the following delivered costs of coal to the project:

- Eastern Region of the Interior Coal Province \$22.00
- Appalachian Region of the Eastern Coal Province \$27.00
- Powder River Region of the Rocky Mountain Coal Province \$ 8.50

7.1.3.2 Materials and Supplies

Factors used were:

- Raw water, \$0.30/1000 gallons
- Operating Supplies, 30% of operating labor

7.1.3.3 Operating Labor

The staffing was estimated for each plant. The operating

labor wage rate used was \$8.00 per hour. The payroll burden was 35% of operating and maintenance labor.

7.1.3.4 Maintenance

Maintenance costs used were:

- Process units, 4% of FCI
- Offsites and Ancillaries, 2% of FCI

7.1.3.5 G & A Overhead

An allowance of 1.5% of the total manufacturing cost was used for the G & A.

7.1.3.6 Property Taxes and Insurance

An allowance of 2.75% of FCI was used for property taxes and insurance.

7.1.4 REQUIRED PRODUCT SELLING PRICE

The RPSP was calculated using the capital investments and operating costs previously described plus the following parameters:

- Design/construction period = 5 years
- Operating period = 20 years
- Total project life = 25 years

- Depreciation = 13 years useful life;
double declining
balance method
- Debt = 65% of total investment,
repaid in equal install-
ments over the 20-year
operating life
- Equity = 35% of total investment
- Interest rate = 9%
- Discounted cash flow (DCF) rate
of return on equity = 12%
- By-product credits:
 - Ammonia = \$140/ton
 - Sulfur = \$ 60/short ton

The RPSP necessary to pay expenses and provide a 12% DCF rate of return on equity was predicted using a computer program. An allowance for inflation for the life of the project was not included in the RPSP model.

7.1.5 POSSIBLE ANNUAL REVENUE

A brief assessment was made of the possible product market values. Market values were assigned to LPG and liquid fuels based on review of published information plus, in some cases, discussion with suppliers and potential users. The values were based on the coal-based fuels being competitive with petroleum based fuels having similar characteristics. Confirmation of the marketability and market values of coal derived fuels is required.

In the case of SNG, the primary evaluation was based on a market value of \$3.25 per thousand cubic feet (MCF). This judgement value is intermediate between intrastate natural gas prices and the value allowed for SNG from naphtha. The true market value for SNG from coal must await commercialization.

The market values used are summarized in Table 7-1; values are presented for four markets consisting of Delaware, Chicago, Houston, and San Francisco.

7.1.6 BENEFIT COST RATIO

Benefit/cost (B/C) ratios were calculated for each high potential process as follows:

$$B/C = \frac{\text{Possible Annual Market Value of Products}}{\text{Required Annual Revenue to Yield 12\% DCF on Equity}}$$

The B/C ratio is presented to rank the processes based on projected profitability levels. When the B/C ratio is greater than one, the projected DCF based on the market value of products is greater than 12%. The higher the B/C ratio, the greater the projected DCF (profitability).

7.2 RESULTS: BASE CASE, PLANT GATE ECONOMICS

The results for projected plant gate economics for the Eastern Region of the Interior Coal Province follow.

7.2.1 FIXED CAPITAL INVESTMENT

The predicted ratios of the fixed capital investments for the four processes are shown in Table 7-2; the relative contributions, on an index basis, of the separate plant sections are also shown. The results indicate that the processes array themselves in the following order of increasing FCIs:

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

7.2.2 TOTAL CAPITAL INVESTMENT

The predicted ratios of total capital investments for the processes are presented in Table 7-3. The relative contributions of the separate contributors to the total investment are given on an index basis.

The ranking of the processes is identical to that obtained for fixed capital investment.

7.2.3 OPERATING COSTS

The ratios of the total operating costs are shown in Table 7-4. Cost contribution by the cost items are shown on an index basis. The processes, ranked in order of increasing operating costs, array themselves as follows:

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

7.2.4 REQUIRED PRODUCT SELLING PRICE (RPSP)

The ratios of total required annual revenues and therefore, unit Btu RPSP are given in Table 7-5. The sub-group contributions are shown on an index basis. The ranking, in order of increasing unit Btu RPSP is:

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

7.2.5 POSSIBLE ANNUAL REVENUE

The ratios of possible annual total revenues for the 600 billion Btu per day of products are presented in Table 7-6. The contribution by the individual products are shown on an index basis. Here the ranking, in decreasing order of possible revenues, is:

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

7.2.6 BENEFIT COST (B/C) RATIO

The ratio of possible annual market revenue to required annual revenue at 12% DCF was calculated for each process, base case. This important ratio is a direct function of projected profitability.

The predicted B/C ratios are:

<u>Process</u>	<u>B/C Ratio</u>
Fischer-Tropsch based	0.88
H-Coal based	0.88
SRC II based	0.85
CSF based	0.77

The results indicate that the B/C ratios for Fischer-Tropsch, H-Coal, and SRC II are approximately equivalent, with CSF showing an approximately 10% lower value. For the project parameters used, all of the process plants are indicated to yield a DCF lower than 12% on equity.

The sensitivity of the B/C ratio to SNG market value is shown in Table 7-7 and Figure 7-1. The sensitivity was projected over a SNG market value range of \$2.25 to \$4.25 per MCF. The results show that the B/C

ratios for the configuration of Fischer-Tropsch used for this study predictably is most sensitive to SNG market value. At a \$4.25/MCF SNG value, the B/C ratio is approximately one. It also indicates that H-Coal and SRC II based plants show similar B/C ratios and similar sensitivity slopes; for CSF the B/C ratio is lower than for H-Coal and SRC II.

7.3 RESULTS: BASE CASE, DELIVERED

A secondary objective of the analysis was to compare the projected technical and economic performances of the high potential processes when delivering fuels to four market areas. To accomplish this, a brief survey of fuel transportation costs was conducted. For reference, the results are summarized in Appendix B.

The first example to be presented is delivery of products from the Eastern Region of the Interior Coal Province to the Chicago marketing area terminal. To illustrate this case, the plant site was arbitrarily conceived to be located in New Athens, Illinois; information is available for this site which was earlier selected for potential use in the government demonstration plant program.

Preferred transportation methods and costs were investigated for movement of all products from New Athens to Chicago. Transport methods investigated included pipelines, barge, and rail. The transportation costs were based on the results of contacts with railroads, pipeline and barge companies, review of available published literature, and inhouse information.

The results indicate that the costs of transporting the products represent the following percentages of the plant gate RPSPs and B/C ratios:

Item	Process			
	SRC II based	H-Coal based	CSF based	Fischer-Tropsch based
Transportation Cost as % of Plant Gate RPSP	4	4	3	6
Ratio of Delivered RPSP	1.00	1.16	1.23	1.46
B/C Ratios, Delivered Basis	0.82	0.84	0.72	0.85

The ranking of the processes is therefore the same based on delivered costs as earlier developed for plant gate costs.

7.4 RESULTS: EFFECT OF ALTERNATE LOCATIONS

An additional secondary objective of the study was to assess the performance and projected economics for the processes in the Appalachian Region of the Eastern Coal Province and the Powder River Region of the Rocky Mountain Province.

Differences which result from operation of the process plants in the three separate regions include coal characteristics, cost of coal, product slate, labor costs, preferred method of supplying the utilities, and others. The effects of these differences were discussed in Section 5.

The effects of the different locations were analyzed. Table 7-8 summarized projected values of FCI ratio, predicted thermal efficiency, and the required feed coal and oxygen rates for the three regions. Table 7-9 shows the relative contributions of the separate accounts to RPSP ratio for the Eastern Coal Province case and Table 7-10 presents similar data for the Rocky Mountain Region.

The summary comparison present in Table 7-11 shows that the RPSP ratios vary significantly between regions, with the Appalachian area the most costly area in which to produce and the Western Rocky Mountain Region the least expensive and giving the SRC II process a greater advantage in this

area. The shift in ratios is mainly attributed to lower coal prices in the West, \$8.25/ton versus \$27.00/ton in the Appalachian Region of the Eastern Coal Province; the western coal used for this study was largely surface mine produced while the Appalachian area used significant underground mining.

Table 7-11 ratios combined with Table E-1 (Appendix E) key cost data indicate lowest potential product costs at the western U.S. site, intermediate costs in the Interior region and somewhat bigger costs in the Appalachian region.

7.4.1 DELIVERED COST, APPALACHIAN AND ROCKY MOUNTAIN REGIONS

Delivery costs for the Appalachian location plant products to the Dover, Delaware marketing center are based on pipeline deliveries. These are approximately double the delivery costs for the base case location to Chicago.

The Western Rocky Mountain Region product delivery to San Francisco, also by pipeline, costs approximately four times that for the base case location to Chicago.

Tabulations of transportation cost percentages and ratios of delivered RPSP are shown below:

Item	Process			
	SRC II based	H-Coal based	CSF based	Fischer-Tropsch based
<u>Appalachian Region to Dover, DE:</u>				
Transportation costs as % of Plant Gate RPSP	9	8	7	5
Ratio of Delivered RPSP	1.00	1.16	1.25	1.37
B/C Ratios, Delivered Basis	0.75	0.77	0.69	0.81
<u>Western Rocky Mountain Region to San Francisco:</u>				
Transportation cost as % of Plant Gate RPSP	22	19	17	18
Ratio of Delivered RPSP	1.00	1.20	1.27	1.41
B/C Ratios Delivered Basis	0.87	0.87	0.79	0.91

Table 7-12 summarizes the projected B/C ratios for plant gate and for delivering of products to the assigned marketing areas. The largest decline in B/C ratios for delivered as compared to plant gate operations occurs in the Western Region.

Table 7-1 - Possible Fuel Market Values
(First Quarter 1978)

Item	Possible Unit Market Value			
	Delaware	Chicago	Houston	San Francisco
SNG (\$/MCF)	3.25	3.25	3.25	3.25
LPG ₅ (\$\$/bbl)	12.00	12.50	12.00	12.00
LPG ₄ (\$/bbl)	13.00	13.50	13.00	13.00
Gasoline (\$/bbl)	17.15	18.00	16.75	17.75
Distillate Fuel Oil No. 2 (\$/bbl)	16.00	15.75	14.50	15.00
Distillate Fuel Oil No. 4 and Heavy (\$/bbl)	14.50	15.00	14.50	14.50
Residual Fuel Oil (\$/bbl)	12.00	12.00	12.00	11.00

Table 7-2 - Ratios of Estimated Fixed Capital Investments

Item	Relative Contribution to FCI			
	SRC II Based	H-Coal Based	CSF Based	Fischer-Tropsch Based
Coal Preparation, Storage, Grinding and Drying	0.06	0.06	0.06	0.05
Process Operations	0.62	0.78	0.91	1.20
Steam and Power Generations	0.09	0.10	0.13	0.15
Offsites and General Facilities	<u>0.12</u>	<u>0.13</u>	<u>0.15</u>	<u>0.10</u>
Total Constructed Cost	0.89	1.07	1.25	1.50
Add: Engineering, Home Office, Fee and Sales Tax @ 12.5%	<u>0.11</u>	<u>0.13</u>	<u>0.15</u>	<u>0.20</u>
Ratio to SRC II	1.00	1.20	1.40	1.70

Table 7-3 - Ratios of Total Capital Investment
Eastern Region of the Interior Coal Province (Base Case)

Item	Relative Contribution to Total Investment			
	SRC II Based	H-Coal Based	CSF Based	Fischer- Tropsch Based
Fixed Capital Investment	0.75	0.91	1.07	1.29
Initial Catalysts	0.01	0.01	0.01	0.02
Start-up Costs	0.05	0.05	0.06	0.07
Interest During Construction	0.10	0.13	0.13	0.15
Commitment Fee	<u>0.02</u>	<u>0.02</u>	<u>0.03</u>	<u>0.03</u>
Total Depreciation Investment	0.93	1.12	1.30	1.56
Working Capital	<u>0.07</u>	<u>0.08</u>	<u>0.10</u>	<u>0.12</u>
Ratio of Total Capital Investment to SRC II	1.00	1.20	1.40	1.68

Table 7-4 - Ratios of Operating Costs

Item	Relative Contribution to Operating Cost			
	SRC II Based	H-Coal Based	CSF Based	Fischer- Tropsch Based
Coal	0.63	0.67	0.73	0.75
Materials, Supplies, and Utilities				
Operating Supplies	0.01	0.01	0.01	0.01
Utilities	0.01	0.01	0.01	0.01
Maintenance Materials and Contract Labor	0.08	0.09	0.11	0.13
Catalysts and Chemicals	<u>0.04</u>	<u>0.07</u>	<u>0.04</u>	<u>0.03</u>
Total Material, Supplies and Utilities	0.14	0.18	0.17	0.18
Labor				
Operating Labor and Supervision	0.02	0.03	0.02	0.02
Maintenance Labor and Supervision	0.04	0.05	0.05	0.07
Payroll Burden	0.02	0.02	0.02	0.03
Plant Overhead/Administration	<u>0.05</u>	<u>0.06</u>	<u>0.06</u>	<u>0.07</u>
Total Labor Costs	0.13	0.16	0.15	0.19
G and A Overhead	0.01	0.01	0.02	0.02
Property Taxes and Insurance	0.09	0.11	0.13	0.15
Ratio to SRC II	<u>1.00</u>	<u>1.13</u>	<u>1.20</u>	<u>1.29</u>

Table 7-5 - Ratios of Required Product Selling Price

Item	Relative Contribution to Unit Btu RPSP			
	SRC II Based	H-Coal Based	CSF Based	Fischer-Tropsch Based
Capital ^a	0.38	0.46	0.54	0.63
Coal	0.42	0.44	0.49	0.49
Operations ^b	0.24	0.31	0.31	0.36
By-product Credit ^c	<u>-0.04</u>	<u>-0.05</u>	<u>-0.06</u>	<u>-0.04</u>
Ratio of RPSP to SRC II	1.00	1.16	1.28	1.44
<p>^a Amortization; 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.</p> <p>^b Labor, utilities, maintenance labor and materials, property taxes, insurance, catalysts and chemicals, supplies, G and A overhead, plant overhead, payroll burden.</p> <p>^c By-product credit based on: sulfur, \$60/ton; ammonia, \$140/ton.</p>				

Table 7-6 - Ratios of Possible Annual Revenues
Eastern Region of Interior Coal Province (Base Case)

Item	Relative Contribution to Annual Revenue			
	SRC II Based	H-Coal Based	CSF Based	Fischer-Tropsch Based
SNG (%)	0.32	0.36	0.30	0.74
LPG ₃	0.04	0.04	0.10	-
LPG ₄	0.02	0.11	0.02	0.12
Gasoline	0.13	0.26	0.27	0.46
No. 2 Fuel	-	-	-	0.03
Heavy Fuel Oil	-	0.04	-	0.02
Residual Fuel Oil	0.44	0.33	0.40	-
Oxygenates	-	-	-	0.07
By-products	<u>0.05</u>	<u>0.06</u>	<u>0.07</u>	<u>0.04</u>
Ratio of Possible Total Revenue to SRC II	1.00	1.20	1.16	1.48

Table 7-7 - Sensitivity of Benefit/Cost Ratio to SNG Market Value
 Eastern Region of the Interior Coal Province (Base Case)

Item	Process			
	SRC II Based	H-Coal Based	CSF Based	Fischer- Tropsch Based
SNG Production as % of Total Fuels, Btu Basis	26	28	25	53
Benefit/Cost Ratio:				
at \$4.25/MCF SNG Price	0.93	0.96	0.83	1.01
at \$3.25/MCF SNG Price (Base Price)	0.85	0.88	0.77	0.88
at \$2.25/MCF SNG Price	0.77	0.80	0.71	0.74

Table 7-8 - Alternate Location Comparison

Process	Eastern Region Interior Coal Province (Base Case)				Appalachian Region Eastern Coal Province				Western Rocky Mountain Region			
	FCI Ratio	Thermal Efficiency HHV Basis (%)	Coal Feed Rate (TPD)	Oxygen Feed Rate (TPD)	FCI ^a (% Change)	Thermal Efficiency (%)	Coal Feed Rate (TPD)	Oxygen Feed Rate (TPD)	FCI ^a (% Change)	Thermal Efficiency (%)	Coal Feed Rate (TPD)	Oxygen Feed Rate (TPD)
SRC 11 Based	1.0	78	30,630	3,145	-2	79	27,873	3,050	+10	71	33,387	3,800
II-Coal Based	1.2	74	32,260	4,420	-1	75	30,324	4,330	+14	60	36,454	5,790
CSF Based	1.4	67	35,710	8,755	-2	68	32,496	8,490	+11	60	38,924	10,770
Fischer-Tropsch Based	1.7	66	36,340	23,600	-2	67	33,069	22,420	+4	59	39,974	25,960
^a FCI, percentage change from base case.												

Table 7-9 - Ratios of RPSP - Appalachian Region of Eastern Coal Province

Item	Relative Contribution to Unit Btu RPSP			
	SRC II	H-Coal	CSF	Fischer-Tropsch
Capital ^a	0.36	0.43	0.51	0.61
Coal	0.45	0.50	0.52	0.53
Operations ^b	0.25	0.29	0.30	0.33
By-product Credit ^c	<u>-0.04</u>	<u>-0.06</u>	<u>-0.06</u>	<u>-0.05</u>
Ratio of RPSP to SRC II	1.00	1.17	1.27	1.42

Table 7-10 - Ratios of RPSP - Western Rocky Mountain Region

Item	Relative Contribution to Unit Btu RPSP			
	SRC II	H-Coal	CSF	Fischer-Tropsch
Capital ^a	0.52	0.65	0.75	0.81
Coal	0.22	0.23	0.25	0.25
Operations ^b	0.32	0.41	0.41	0.46
By-product Credit ^c	<u>-0.06</u>	<u>-0.06</u>	<u>-0.08</u>	<u>-0.06</u>
Ratio of RPSP to SRC II	1.00	1.23	1.33	1.46

^aDepreciation; borrowing cost, 65% of the total investment borrowed at 9% interest, with principal repaid in equal installments over the 20-year project term; working on funds not yet drawn; 12% return on equity funds; income tax.

^bLabor, utilities, maintenance labor and materials, property taxes, insurance, catalysts and chemicals, supplies, G&A overhead, plant overhead, payroll burden.

^cBy-product credit based on: sulfur, \$60/ton; ammonia, \$140/ton.

Table 7-11 - Regional Comparison of RPSP Ratios

Region	Base Process			
	SRC II	H-Coal	CSF	Fischer-Tropsch
Eastern (Base Case)	1.00	1.16	1.28	1.44
Appalachian	1.04	1.22	1.32	1.48
Western	0.81	0.99	1.18	1.07

Table 7-12 - Regional Comparison of B/C Ratios

Region	Cost Type	Base Process			
		SRC II	H-Coal	CSF	Fischer-Tropsch
Eastern (Base Case)	Plant Gate	0.85	0.88	0.77	0.88
	Delivered	0.82	0.84	0.72	0.85
Appalachian	Plant Gate	0.82	0.83	0.74	0.85
	Delivered	0.75	0.77	0.69	0.81
Western	Plant Gate	1.06	1.03	0.92	1.07
	Delivered	0.87	0.87	0.79	0.91

Figure 7-1 - Benefit/Cost Ratio Sensitivity to SNG Market Value

SECTION 8
DISCUSSION

SECTION 8

DISCUSSION

8.1 INTRODUCTION

The Section 7 economic comparison indicates that, for the process configurations, product compositions and economic assessment procedures used, the SRC II based technology offers the potential for lower fixed capital investments (see Table 7-2) and lower RPSP (see Table 7-5). The group of high potential processes show the following ranking in order of increasing capital cost:

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

When a benefit/cost (B/C) ratio index is applied, SRC II based, H-Coal based, and Fischer-Tropsch based processes show similar B/C's, with CSF indicating a lower B/C. Since the B/C ratio is defined as:

$$B/C = \frac{\text{Possible Annual Product Revenue}}{\text{Required Annual Revenue @ 12\% DCF on Equity,}}$$

The comparison of B/C and RPSP confirm Table 7-7 information that the H-Coal and Fischer-Tropsch configurations used in this study produce higher potential valued product slates than the SRC II based configuration.

The Fischer-Tropsch products differ from those produced by SRC II and H-Coal in that they have nil sulfur, nitrogen, and particulate contents. They also tend to be higher in aliphatic rather than aromatic in composition and would therefore have potential marketability as petrochemical feedstocks.

To summarize, the Fischer-Tropsch products appear to have certain advantages relative to those produced in the hydroliquefaction processes (SRC II and H-Coal) with regard to environmental acceptability and potential value to the petrochemical industry.

Further suggestions for interpretation of the Section 7 comparative economics are discussed below and additional analyses presented.

The comparative economics developed in this study represent the results of a limited effort analysis. During the course of this limited effort, the characteristics of some 33 coal liquefaction processes were examined and four were selected for more detailed comparisons. Because adequate detail was not available for the Exxon Donor Solvent (EDS) process from DOE at the start of the study, DOE agreed that it would be excluded. Similar future studies should include the EDS, as well as the Mobil-M process.

In retrospect, the question of whether significant additional work (manhours) would seriously alter the results was briefly examined. To do this, more comprehensive conceptual design/economic evaluations by major contractors for SRC II based, H-Coal based, CSF based, and Fischer-Tropsch based process plants were reviewed, their results were compiled and analyzed, and the results are summarized later in this report section.

Another question retrospectively evaluated was the effect of differences in product slate on the comparative economics. A first approach to this analysis was a comparison of the hydrogen-to-carbon (H/C) ratio with fixed capital investment and required product selling price. The rationale behind this was that liquefaction involves the addition of hydrogen to coal which has a carbon hydrogen composition of approximately $\text{CH}_{0.8}$. Product aromatics will have a C:H composition such as CH, while aliphatics will approach CH_2 and SNG is CH_4 . As more hydrogen is added to the coal, the products tend to be "more liquid," have lower sulfur and nitrogen contents, and tend to have a higher market value. The question of whether product H/C ratios might provide some indicator of the differences in economics to be expected between process plants which produce differing product slates was therefore briefly examined and the results are summarized below.

8.2 SUMMARY: MAJOR CONTRACTORS' CONCEPTUAL DESIGN/ECONOMIC EVALUATION RESULTS

The objective of this brief survey was to determine if similar economic comparison conclusions might have been drawn from inspection of the results contained in conceptual design/economic evaluation reports which used significantly more effort (manhours). To do this, reports for SRC II (Oil/Gas),⁶ H-Coal),¹² CSF,¹⁴ and Fischer-Tropsch⁵ plants were reviewed. Similar comprehensive design reports were not available to us for the M-Gasoline or EDS processes. For the purpose of this comparison, captive coal mines and coal preparation costs were deleted from the reference report estimates. Also, the capacity of each plant was adjusted to 670 billion Btu/day of products (the Oil/Gas plant output which was the higher of the group) and the estimates escalated to first quarter 1978 dollars. No adjustments were made to process configurations or efficiencies.

The product slates for the four processes, expressed as percent of total fuels produced, on a Btu basis, are shown in Table 8-1. The results indicate that product slates vary widely making valid comparisons difficult. Note that the CSF process produces a very high percentage of gasoline (by extensive hydrocracking of heavy liquids at a high capital cost). On the other hand, the H-Coal process included very little upgrading; further, while gas is made by the H-Coal process none is shown as product indicating a designers option to consume this product to satisfy plant fuel requirements.

The fixed capital investment ratios are shown in Table 8-2; for comparison, the ratios developed in this study and presented in Table 7-2 are also presented. Only limited conclusions can be safely drawn from this comparison. These conclusions appear to be that when a given contractor compared several processes, the processes arrayed themselves in the same FCI ranking as developed in this study. To illustrate, the SRC II based and Fischer-Tropsch based conceptual designs/economic evaluations were developed by the same contractor, as were the H-Coal based and CSF based.

To properly compare processes where the designs/economic evaluations were developed by different contractors would require adjustments to factors such as thermal efficiency, and others.

To summarize, review of major contractors' conceptual designs/economic evaluations tends to corroborate the FCI rankings developed in this study. Conversely, comprehensive designs/economic evaluations which contradict these rankings are not known to exist at this time.

8.3 EFFECT OF PRODUCT COMPOSITION ON ECONOMICS

An objective of any comparison of process potential is to compare the economics for production of identical product compositions. Pragmatically, for a limited effort study such as reported here, this objective can be approached but not achieved. As an alternate, it would be valuable to define, in a simple manner, the economic impact of differing product compositions. The availability of such a correlation would assist in making meaningful comparisons between processes which produce different products. A brief start was made at the end of this study to better understand the relationship between product composition and economics.

A first approach was to investigate the relationship between the H/C ratio (weight basis), FCI ratio, and RPSP ratio. The incentive to look at this simple ratio is that as it increases, plant complexity increases, heteroatom content (S,N) decreases, and the products tend to be more "liquid."

To supplement the information from the high potential processes from this study, a brief arbitrary assessment of procedure and possible economics for production of SNG, using Fischer-Tropsch based technology, and 95% H₂ technology,⁵ was developed. This second-order supplementary information is included as Appendix C.

H/C ratios and FCIs for plants to produce 600 billion Btu of fuels per day are shown in Table 8-3 and Figure 8-1. For SRC II based, H-Coal based, and CSF based, the H/C ratio increases in the same ranking as the

FCI ratio. Also shown in Figure 8-2 are H/C and FCI ratios for a modified SRC plant in which incremental hydrotreating capacity is added;²⁰ while the economic assessment bases differed from those used in this study, these independent predictions also show the same trend.

Figure 8-3 presents relative RPSP as a function of product H/C ratio. Here again a trend consistent with the prior rankings of the hydroliquefaction processes SRC II based, H-Coal based, and CSF based is evident.

With some judgment, the results indicate a different relationship for indirect liquefaction processes such as Fischer-Tropsch and SNG. It appears that the cost of complete gasification of the coal to produce syngas dominates the economics and the relative fixed capital investments are similar for SNG and Fischer-Tropsch.

For approximate comparative purposes, Figure 8-4, a least squares fit from plotted data, indicates product market value range as a function of H/C ratio. Here again, an upward trend of market value appears as the H/C ratio increases. It is of interest that a selling price for SNG consistent with the other fuels considered would be \$3.75 per million Btu, a value intermediate between the base and high values used in the Benefit/Cost ratio comparison.

In summary, Figures 8-1 and 8-2 indicate that production of products with increasing hydrogen contents, plus the companion changes in sulfur content and product composition, can be expected to be accomplished at higher costs. These trend results, certainly not highly quantitative at this stage of development, highlight the need to compare process capabilities using equal product compositions. It also begs the question whether there are significant differences between major process contenders if they produce the same products. The answer to this question should recognize that the primary liquefaction facilities may represent approximately 25% of the fixed capital investment with the other 75% consisting of coal preparation, product recovery, and product refining. The trends illustrated in Figures 8-1 and 8-2 are considered significant in any comparison of liquefaction

technologies and can be used as a guide to approximate economic impact of differing product slates for the appropriate process type.

Table 8-1 - Summary of Product Slates from
Reference ^{5,6,12,14} Conceptual Designs

Product (% of Total Fuels, Btu Basis)	Process			
	Oil/Gas ⁶ (SRC II)	H-Coal ¹²	CSF ¹⁴	Fischer- ⁵ Tropsch
SNG	26.0	a	17.0	53.0
LPG	6.0		13.0	3.0
Gasoline			70.0	
Naphtha	8.0	38.0		19.0
Diesel Fuel				17.0
Heavy Fuel Oil				6.0
Syncrude		62.0		
Residual Fuel Oil	60.0			
Oxygenates	_____	_____	_____	<u>2.0</u>
Total	100.0	100.0	100.0	100.0
^a SNG is produced but is consumed in process as fuel.				

Table 8-2 - Fixed Capital Investment Ratios
from Reference 5,6,12,14 Conceptual Designs
(Basis: 670 Billion Btu Products/Day)

Item	Process			
	Oil/Gas ⁶ (SRC-II)	H-Coal ¹²	CSF ¹⁴	Fischer- ⁵ Tropsch
FCI ratio from reference reports, 670 billion Btu/day	1.0	1.7	1.9	1.6
FCI ratio, this study, Table 7-2	1.0	1.2	1.4	1.7

Table 8-3 - Plant Product H/C and FCI Ratios

Process	H/C	Fixed Capital Investment
SRC II based	0.128	1.0
H-Coal based	0.150	1.2
CSF based	0.162	1.4
F-T based	0.284	1.7
SNG (F-T based)	0.323	1.7
H ₂	4.29	1.4

Figure 8-1 - Fixed Capital Investment Ratio
Vs. Average Hydrogen To Carbon Weight Ratio

**Figure 8-2 - Fixed Capital Investment Ratio and Oil/Gas Desulfurization
Vs. Oil/Gas Products Hydrogen To Carbon Weight Ratio**

Figure 8-3 - Required Product Selling Price
Vs. Average Hydrogen to Carbon Weight Ratio

Figure 8-4 - Product Market Value
Vs. Hydrogen to Carbon Weight Ratio

SECTION 9
LITERATURE CITED

SECTION 9

LITERATURE CITED

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APPENDIX A
EVALUATION PROCEDURES AND RANKINGS

APPENDIX A

EVALUATION PROCEDURES AND RANKINGS

A.1 INTRODUCTION

This appendix presents preliminary rankings of the commercial potentials for candidate processes and the procedures used to develop these rankings.

The scope of the study summarized in this report was broad, incorporating the total field of coal liquefaction. The resource application, defined by contract period and level of funding, required a direct approach to define highest potential processes early in the study in order to permit maximum time/effort to develop preferred product slate, processing procedures, projected economics, and opinions regarding expected performance. To aid in the high potential process selection program, procedures were defined to develop numerical ratings of the separate process characteristics, to weight the ratings, and then, to add these weighted ratings to produce projected rankings of the processes.

The number of processes considered, 33, precluded detailed and quantitative analyses of all characteristics of each process at the ranking stage of this study. What was done was to develop defined techniques to compare 15 different characteristics of each candidate process and apply best engineering and economic assessment judgement to the data and information known for each candidate. The goal was to be objective; in a practical sense, because of resource limitations, certain elements had to be judgementally subjective.

The results of this ranking effort may be viewed as the opinions of a particular group at a given point in time based on nonproprietary information available to them at that time. A significant amount of analytical information is presented which represents a contribution to the coal liquefaction development program.

The rankings developed represent one input to the high potential process selection decision. The rankings were viewed against the perspective gained during work consisting of approximately 130 manyears of work in the field prior to recommending a list of high potential processes for more detailed study.

The ranking techniques and weighting factors used were independently developed for use in this study. They used the experience of similar studies for other fields²¹ as background.

A.2 PROCEDURES

Two separate ranking categories were used; they were:

- Development Status
- Commercial Potential

A total of fifteen separate process characteristics, as listed in Table A-1, were defined and evaluated under these two ranking headings. The weighted rating of each of these characteristics were added to obtain sums used as a guide to process ranking.

Details of the ranking procedures used are presented in the following paragraphs.

A.2.1 DEVELOPMENT STATUS

The Development Status Ranking was composed of three factors with weighting as follows:

<u>FACTORS</u>	<u>WEIGHT</u>
D1; Technology Status	3 1/3
D2; Equipment Development Requirements	3 1/3
D3; Process Development Status	<u>3 1/3</u>
Total Weight	10.0

The numerical values assigned to the individual ratings and weighting factors were based on analysis and judgement. Rating values selected ranged from 0 to 10 and weighting factors from 0 to 10, adjusted so that the maximum possible weighted ranking score was 100.

Items comprising each of the three factors and guidelines for establishing the ratings were:

A.2.1.1 D1; Technology Status (weight 3 1/3)

<u>Rating</u>	<u>Status</u>
1	Laboratory work not begun
2	Bench scale/laboratory
3	PDU work underway
5	Pilot plant data available
7	Demonstration plant engineering complete
10	Commercial plant in operation

Where the following definitions were applied:

<u>Operation Type</u>	<u>Coal Feed</u>
Bench scale/laboratory	less than 1 TPD
PDU	1 to 5 TPD
Pilot Plant	6 to 100 TPD
Demonstration plant	101 to 3000 TPD
Commercial size plant	more than 3000 TPD

A.2.1.2 D2; Equipment Development Requirements (weight 3 1/3)

Equipment development is defined as the summation of twice the number of new and complete items of equipment that are not presently available, plus the number of items of equipment that require extension of existing technology to new and more severe conditions.

<u>Rating</u>	<u>Development Requirements</u>
0	7
1	6
2	5
4	4
5	3
6	2
8	1
10	0

Examples of equipment not presently available:

- Indirect slurry heaters
- Centrifugal slurry pumps above 100 psig discharge pressure
- Lock hoppers operating above 500 psig

- Coal extruders discharging above 500 psig
- Rotary filters operating above 300°F

Examples of equipment requiring an extension of existing technology:

- Rotating fluidized bed

A.2.1.3 D3; Process Development Status (weight 3 1/3)

<u>Rating</u>	<u>Evaluation</u>
1	Critical steps unproven at any scale, assumptions made.
2	Critical process steps unproven at any scale, but theoretically sound.
3	All process steps proven at bench scale.
5	Critical process steps proven at PDU scale; difficult but solvable problems exist.
7	Critical process steps proven at pilot plant scale; difficult but solvable problems exist.
10	All process steps proven in an integrated demonstration plant; none or only minor problems exist.

Examples of difficult but solvable problems:

- Flashing (one or more) in presence of solids
- Fractionation in presence of solids

A.2.2 COMMERCIAL POTENTIAL

The Liquefaction Potential Ranking is composed of twelve factors as listed below:

<u>Ranking Factor</u>	<u>Weighting Factor</u>
P1; Complexity	0.855
P2; Severity	0.945
P3; Industrial Capacity	0.450
P4; Feed Versatility	0.675
P5; Product Versatility	0.675
P6; Liquefaction Efficiency	0.900
P7; Sulfur Reduction Efficiency	1.575
P8; Conversion Efficiency	0.540
P9; Product Premium	1.260
P10; Product Compatibility	0.180
P11; Product Hazard	0.945
P12; Uncertainty Factor	1.000
Total Weight	10.000

The numerical values assigned to the individual ratings and weighting factors are based on analysis and judgement. Ratings have been selected to have values ranging from 0 to 10, with the weighting factors adjusted so that the maximum possible ranking score is 100.

The ratings used for the separate factors and the guidelines defined for the process assessors are summarized in the following paragraphs.

A.2.2.1 Pl; Process Complexity (weight 0.855)

Complexity is the sum of Internal and External Factors which are defined as follows:

- Internal Factors Value

Number of independent reaction stages	1 each
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Number of principle recycle streams	1 each
-------------------------------------	--------

- External Factors

Number of process units that feed the liquefaction plant e.g.

<u>Process Unit</u>	<u>Value</u>
Coal drying	1
Hydrogen by coal oxidation	2
Methane steam reformer	1
Oxygen plant	0
Coal grinding	0
Steam generation	0

The Complexity Rating is the numerical sum of the above value items:

<u>Complexity</u>	<u>Rating</u>
10	0
9	1
8	2
7	3

6	4
5	5
4	6
3	7
2	8
1	9
0	10

A.2.2.2 P2; Process Severity (weight 0.945)

Operating pressures and temperatures were categorized as follows:

<u>Category</u>	<u>Pressure (psia)</u>	<u>Temperature (°F)</u>
High	>1000	>1000
Medium	501-1000	651-1000
Low	< 500	< 650

Each process was rated for severity according to the following:

<u>Pressure</u>	<u>Temperature</u>		
	<u>High</u>	<u>Medium</u>	<u>Low</u>
High	1	3	6
Medium	2	5	8
Low	4	7	10

^aFor pressures over 3000 psi, at any temperature, rating = 1.

A.2.2.3 P3; Industrial Capacity (weight 0.45)

Industrial capacity was defined to be a function of the maximum amount of liquid product, expressed in BPSD, that can be produced

from one primary coal conversion reactor. The maximum practical reactor size for high pressure operation was defined as a vessel 12-foot ID with an internal volume not to exceed 8,000 feet.

The relationship between industrial capacity rating and reactor capacity as defined above is shown in Figure A-1.

A.2.2.4 P4; Feed Versatility (weight 0.675)

Each process is rated according to its ability to liquefy the following four basic types of coal, with a liquid yield variation of less than 25%.

1. Low-volatile bituminous coal
2. Med- and High-volatile bituminous coal
3. Sub-bituminous coal
4. Lignite

Types 1. and 2. are considered to be agglomerating coals.

<u>Versatility</u>	<u>Rating</u>
Process handles 1 coal	2.5
Process handles 2 coals	5.0
Process handles 3 coals	7.5
Process handles 4 coals	10.0

A.2.2.5 P5; Product Versatility (weight 0.675)

<u>Product State Versatility</u>	<u>Rating</u>
Only one liquid product, cannot be varied by operational or catalyst changes.	1
Either one of two liquid products can be produced.	2

A multi-liquid product slate is produced, 4
and cannot be varied by operational or
catalyst changes.

A multi-liquid product slate is produced. 6
Preliminary product can be shifted to next
higher or lower boiler range product.

A multi-liquid product slate is produced. 8
Primary product can be shifted to next
two higher or lower boiler range products.

Any primary product between heavy fuel oil 10
and LPG can be produced.

A.2.2.6 P6; Liquefaction Efficiency (weight 0.900)

Liquefaction efficiency is defined as the percentage
of the higher heating value of the coal feed plus added hydrogen coal equi-
valent recovered as liquid hydrocarbons, excluding liquified petroleum gases
(LPGs).

$$E_L = \frac{\text{HHV (liquids with BP LPG)}}{\text{HHV (Coal Feed + H}_2 \text{ CE)}}$$

$$\text{H}_2 \text{ CE} = 0.7 (\text{HHV}_{\text{H}_2} \text{ consumed})$$

where: E_L = Liquefaction Efficiency

$\text{H}_2 \text{ CE}$ = Coal Equivalent of H_2 consumed

<u>Liquefaction Efficiency</u>	<u>Rating</u>
0%	0
50%	5
100%	10

A.2.2.7 P7; Sulfur Reduction Efficiency (weight 1.575)

Sulfur reduction efficiency is the difference between coal and product oil sulfur contents, measured as lbs. of sulfur per MM Btu, expressed as a percentage of the coal sulfur content.

$$E_{SR} = \frac{S_c - S_o}{S_c} \times 100$$

where: E_{SR} = Sulfur Reduction Efficiency

S_c = Lb. Sulfur in Coal per MM HHV

S_o = Lb. Sulfur in Product Oil per
MM Btu HHV

<u>Sulfur Reduction Efficiency</u>	<u>Rating</u>
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0%	0
50%	5
100%	10

A.2.2.8 P8; Conversion Efficiency (weight 0.540)

Conversion efficiency is a measure of the degree of coal liquefaction achieved; it is defined as 100 minus the percentage of C_5 + hydrocarbons (Btu basis) boiling above 850°F which fail to meet a Bunker Fuel Oil viscosity specification of 180 SSF at 122°F. This may be expressed as follows:

Conversion Efficiency = 100 - % C_5 , + with BP 850°F, 180 SSF @ 122°F.

<u>Conversion Efficiency</u>	<u>Rating</u>
------------------------------	---------------

0%	0
50%	5
100%	10

A.2.2.9 P9; Product Market Value (weight 1.260)

Product market value is defined as the anticipated market value of the coal derived liquids, expressed as a percentage of the value of an equivalent petroleum based fuel, when due consideration is given to fuel octane or cetane number, volatility, heating value, pour point, and carbon to hydrogen ratio.

The overall product market values for each process were determined by dividing the wide boiling range coal liquid into cuts of equivalent petroleum based fuels, according to ASTM specifications. Product market values for each cut were then assessed and the overall market value obtained by weighing the individual values on a Btu basis.

<u>Product Market Value as % of Petroleum Based Product</u>	<u>Rating</u>
0%	0
50%	5
100%	10

A.2.2.10 P10; Product Compatibility (weight 0.180)

Product compatibility is a measure of the ability of the coal derived liquids to maintain satisfactory storage stability in the presence of equivalent petroleum based fuels. The significance of this factor is that if a coal oil fuel is not compatible with an equivalent petroleum fuel, then users would be required to dedicate tankage specifically for coal based fuels.

Product compatibility is defined as the percentage of coal oil products, on a Btu basis, that are fully compatible with the equivalent petroleum based product. Since compatibility information for many processes is inadequate, ratings have been made on the general assumption that only residual fuels, excluding those produced by reaction of synthesis

gas, i.e., a mixture of $\text{CO} + \text{H}_2$, would be incompatible with a petroleum base fuel.

<u>Compatibility</u>	<u>Rating</u>
0%	0
50%	5
100%	10

A.2.2.11 P11; Product Hazard (weight 0.945)

The potential hazard of coal oil fuels to the safety and health of humans, when encountered under normal conditions of use, is related to toxic and carcinogenic properties of the fuel.

Several classification schemes were reviewed before establishing the rating guidelines. In one, Sax, I. N., Dangerous Properties of Industrial Materials, Van Nostrand Reinhold Co., 1975, the following system of toxicity rating has been used to indicate the relative hazard of these fuels as applied to human exposure.

U	=	Unknown
0	=	No toxicity
1	=	Slight toxicity
2	=	Moderate toxicity
3	=	Severe toxicity

The acute toxicity ratings, referring to a single exposure of durations up to several hours, have been used in this evaluation.

Another procedure reviewed for expressing toxic levels is the Threshold Limit Value (TLV), which is pertinent for industrial and occupational exposure restrictions, and refers to airborne concentrations, at which it is believed nearly all workers may be repeatedly exposed for an 8 hour day, 5 days a week. Substances carcinogenic for humans may be

divided into those substances with an assigned TLV, and those without an assigned TLV.

Since toxicity data for most liquefaction products is meager, the hazard rating has been based upon published data for similar substances, e.g., coal tar fuels, pure aromatics, and petroleum fuels.

Product hazard ratings for the individual product cuts were weighted to obtain the overall product hazard rating.

The product hazard rating is determined according to the following guidelines:

<u>Product Hazard</u>	<u>Rating</u>
Extremely Toxic <ul style="list-style-type: none">● carcinogenic with no TLV	0
Highly Toxic <ul style="list-style-type: none">● carcinogenic with TLV 1 ppm V	2
Very Toxic <ul style="list-style-type: none">● acute toxicity rating = 3 or U● carcinogenic with TLV 10 ppm	4
Moderately Toxic <ul style="list-style-type: none">● acute toxicity rating = 2● carcinogenic with TLV 200 ppm	6
Slightly Toxic <ul style="list-style-type: none">● acute toxicity rating = 1● carcinogenic with TLV 500 ppm	8
Harmless <ul style="list-style-type: none">● acute toxicity rating = 0● non carcinogenic	10

A.2.2.12 P12; Uncertainty Factor (weight 1.0)

This factor, which reflects the degree of confidence in the published data is taken to be proportional to the process Development Status Ranking, defined on Page 4-4.

<u>Development Status Ranking</u>	<u>Rating</u>
0%	0
50%	5
100%	10

A.3 NUMERICAL RANKINGS

A summary of the Development Status and Commercial Potential ratings for the thirty-three processes is shown in Table A-2. Table A-3 presents the unadjusted Development Status rankings for the processes while Table A-4 presents similar rankings for the Commercial Potential category.

Details of the ratings for the separate factors in the Development Status Rating are presented in Table A-5, and similar details for the Commercial Potential Rating is shown in Table A-6.

A.4 DISCUSSION OF RANKINGS

The ranking scheme used has been designed to focus attention on those coal liquefaction processes which appear most promising for future liquid fuels production.

Interpretation of the results of this type of analysis requires judgment since the additive system of rating used can give misleading results when extreme values of certain evaluation factors are encountered, e.g. the impact of inclusion of a small number of extreme values must be assessed.

A.4.1 DEVELOPMENT STATUS

The Development Status rating is a measure of how far a process has progressed towards full commercialization, given a cost-development time relationship similar to that found in the process industries. Accelerated funding of a particular process would have the effect of reducing the time required for commercial development.

In general, the following interpretations of the ratings are suggested:

- Development Status rating, 90

Processes in this category are considered to be either commercial or at the commercial prototype stage.

- Development Status rating, 50 to 90

Processes in this category are considered to be capable of demonstrating commercial operation by the year 1990 if due diligence is used in pursuing their development.

- Development Status rating, 40 to 50

Processes in this category are considered to be able to demonstrate commercial operation later than 1990, but before the year 2000; again based on use of due diligence.

- Development Status rating, 40

Processes in this category are considered to be able to demonstrate commercial operation only after the year 2000.

A.4.2 COMMERCIAL POTENTIAL

A review of the commercial potential ranking indicates that 12 coal conversion processes scored points 60 or better, and seven other processes received scores ranging from 55 up to 59 points; these are considered candidates.

The processes with scores of 60 or more consist of:

- 8 indirect processes
- 3 catalytic hydroliquefaction processes
- 1 donor solvent process

The economics and potential of the indirect processes are heavily dependent on the gasification process used to produce syngas. Complete gasification of feed coal to syngas, in general, leads to relatively high unit production costs.

The catalytic hydroliquefaction processes offer good potential and should be capable of demonstrating commercial potential by 1990 if due diligence is used. Catalyst performance and cost per unit of production remain a question.

The seven processes in the group scoring 55 to 59 points are in varying stages of development. Three of the seven processes are not considered to be capable of commercial demonstration by the year 2000 while the SRC I, SRC II and Exxon Donor Solvent should, with due diligence, be capable of demonstrating commercial potential by 1990. One of these processes, 2.3.2 SRT Coal Hydrolysis, is a low yield high quality liquid producer and should therefore, probably be excluded from this evaluation.

The remaining processes received scores below 55, and are considered less attractive candidates for future commercial liquefaction application, although possibly having future application for coal gasification, or specialty chemicals production (e.g., metallurgical coke).

An overview of the potential rankings shows that the most promising processes for liquefaction are based upon hydrol liquefaction (processes identified by the prefix 1.0) and indirect (processes identified by the prefix 3.0) technology.

Table A-1 - Evaluation Criteria

Ranking Category	Characteristics	Description of Characteristics
Development Status	3	<ol style="list-style-type: none"> 1. Technology status 2. Equipment development requirements 3. Process development status
Commercial Potential	12	<ol style="list-style-type: none"> 1. Process complexity 2. Process severity 3. Industrial capacity 4. Feed versatility 5. Product versatility 6. Liquefaction efficiency 7. Sulfur reduction efficiency 8. Conversion efficiency 9. Product premium 10. Product compatibility 11. Product hazard 12. Uncertainty factor

Table A-2 - Summary of Development Status and
Commercial Potential Ratings

Process Number	Process	Development Status Rating	Commercial Potential Rating
1.1.1	Solvent Refined Coal (SRC) I	53.3	55.1
1.1.1	Solvent Refined Coal (SRC) II	53.3	56.5
1.1.2	UOP Extraction Process	36.7	52.5
1.2.1	Gulf Catalytic Coal Liquids (CCL)	43.3	63.0
1.2.2	SYNTHOIL	43.3	64.3
1.2.3	H-Coal	50.0	62.6
1.2.4	Clean Fuel from Coal (CFC)	36.7	55.6
1.2.5	Conoco Zinc Halide Hydrocracking	30.0	57.6
1.3.1	Exxon Donor Solvent (EDS)	46.7	56.1
1.3.2	ADL Extractive Coking	43.3	53.5
1.3.3	Consol Synthetic Fuel (CSF)	53.3	63.6
2.1.1	Char-Oil-Energy-Development (COED), RMP	70.0	46.5
2.1.2	Char-Oil-Energy-Development (COED), FMC	73.3	42.9
2.1.3	Occidental Research Corp. Coal Flash Pyrolysis	60.0	46.3
2.1.4	TOSCOAL	73.3	52.6
2.1.5	Lurgi-Ruhrgas		52.5
2.2.1	U.S. Steel Clean Coke	40.0	48.3
2.2.2	Coalcon	53.3	55.6
2.3.1	BNL Rotating Fluidized Bed	13.3	56.4
2.3.2	Short Residence Time (SRT) Coal Hydro- pyrolysis	36.7	55.6
2.3.3	Intermediate Coal Hydrogenation	33.3	52.7
2.3.4	Schroeder's Rapid Hydrogenation	36.7	54.4
2.3.5	BNL Flash Hydropyrolysis	30.0	50.1
2.3.6	Rockwell International Direct Coal Hydrogenation	33.3	46.6
3.1.1	Flame Sprayed Catalyst (FSC)	50.0	74.3
3.1.2	ARGE (Arbeit Germeinschaft, Ruhrchemie/Lurgi)	100.0	78.2
3.1.3	SYNTHOIL	100.0	75.6
3.2.1	High Pressure Methanol Process	100.0	67.3
3.2.2	ICI LP/LT Methanol	100.0	68.6
3.2.3	Lurgi LP/LT Methanol Synthesis	100.0	69.5
3.2.4	Three-Phase Methanol Synthesis	50.0	62.7
3.3.1	M-Gasoline (Mobil)	50.0	67.7
4.1.1	Supercritical Gas Extraction (SCE)	36.7	37.5

Table A-3 - Development Status Process Rankings

Process Number	Process	Development Status Rating
3.1.2	ARGE (Arbeit Germeinschaft, Ruhrchemie/Lurgi	100.0
3.1.3	SYNTHOIL	100.0
3.2.1	High Pressure Methanol Process	100.0
3.2.2	ICI LP/LT Methanol	100.0
3.2.3	Lurgi LP/LT Methanol	100.0
2.1.5	Lurgi-Ruhrgas	96.7
2.1.2	Char-Oil-Energy-Development (COED), FMC	73.3
2.1.4	TOSCOAL	73.3
2.1.1	Char-Oil-Energy-Development (COED), RMP	70.0
2.1.3	Occidental Research Corp. Coal Flash Pyrolysis	60.0
1.1.1	Solvent Refined Coal (SRC) I	53.3
1.1.1	Solvent Refined Coal (SRC) II	53.3
1.3.3	Consol Synthetic Fuel (CSF)	53.3
2.2.2	Coalcon	53.3
1.2.3	H-Coal	50.0
3.1.1	Flame Sprayed Catalyst (FSC)	50.0
3.2.4	Three Phase Methanol Synthesis	50.0
3.3.1	M-Gasoline (Mobil)	50.0
1.3.1	Exxon Donor Solvent (EDS)	46.7
1.2.1	Gulf Catalytic Coal Liquids (CCL)	43.3
1.2.2	SYNTHOIL	43.3
1.3.2	ADL Extractive Coking	43.3
2.2.1	U.S. Steel Clean Coke	40.0
1.1.2	UOP Extraction Process	36.7
1.2.4	Clean Fuel from Coal (CFFC)	36.7
2.3.2	Short Residence Time SRT) Coal Hydropyrolysis	36.7
2.3.4	Schroeder's Rapid Hydrogenation	36.7
4.1.1	Supercritical Gas Extraction (SCE)	36.7
2.3.3	Intermediate Coal Hydrogenation	33.3
2.3.6	Rockwell International Direct Coal Hydrogenation	33.3
1.2.5	Conoco Zinc Halide Hydrocracking	30.0
2.3.5	BNL Flash Hydropyrolysis	30.0
2.3.1	BNL Rotating Fluidized Bed	13.3

Table A-4 - Commercial Potential Process Rankings

Rank	Process Number	Process	Commercial Potential Rating
1	3.1.2	ARGE (Arbeit Germeinschaft, Ruhrchemie/Lurgi)	78.2
2	3.1.3	SYNTHOL	75.6
3	3.1.1	Flame Sprayed Catalyst (FSC)	74.3
4	3.2.3	Lurgi LP/LT Methanol	69.5
5	3.2.2	ICI LP/LT Methanol	68.6
6	3.3.1	M-Gasoline (Mobil)	67.7
7	3.2.1	High Pressure Methanol Process	67.3
8	1.2.2	SYNTHOIL	64.3
9	1.3.3	Consol Synthetic Fuel (CSF)	63.6
10	1.2.1	Gulf Catalytic Coal Liquids (CCL)	63.0
11	3.2.4	Three-Phase Methanol Synthesis	62.7
12	1.2.3	H-Coal	62.6
13	1.2.5	Conoco Zinc Halide Hydrocracking	57.6
14	1.3.1	Exxon Donor Solvent (EDS)	56.8
15	1.1.1	Solvent Refined Coal (SRC) II	56.5
16	2.3.1	BNL Rotating Fluidized Bed	56.4
17	1.2.4	Clean Fuel from Coal (CFFC)	55.6
18	2.2.2	Coalcon	55.6
19	2.3.2	Short Residence Time (SRT) Coal Hydropyrolysis	55.6
20	1.1.1	Solvent Refined Coal (SRC) I	55.1
21	2.3.4	Schroeder's Rapid Hydrogenation	54.4
22	1.3.2	ADL Extractive Coking	53.5
23	2.3.3	Intermediate Coal Hydrogenation	52.7
24	2.1.4	TOSCOAL	52.6
25	1.1.2	UOP Extraction Process	52.5
26	2.1.5	Lurgi-Ruhrgas	52.5
27	2.3.5	BNL Flash Hydropyrolysis	50.1
28	2.2.1	U.S. Steel Clean Coke	48.3
29	2.3.6	Rockwell International Direct Coal Hydrogenation	46.6
30	2.1.1	Char-Oil-Energy-Development (COED), RMP	46.5
31	2.1.3	Occidental Research Corp. Coal Flash Pyrolysis	46.3
32	2.1.2	Char-Oil-Energy-Development (COED), FMC	42.9
33	4.1.1	Supercritical Gas Extraction (SCE)	37.3

Table A-5 - Development Status Rating

Process Number	Process	Technology Status ^a D1	Equipment Development Requirements ^a D2	Process Development Status ^a D3	Score (Max = 100)
1.1.1	Solvent Refined Coal (SRC) I	5	5	6	53.3
1.1.1	Solvent Refined Coal (SRC) II	5	5	6	53.3
1.1.2	UOP Extraction Process	3	5	3	36.7
1.2.1	Gulf Catalytic Coal Liquids	3	5	5	43.3
1.2.2	SYNTHOIL	3	5	5	43.3
1.2.3	H-Coal	5	4	6	50.0
1.2.4	Clean Fuel from Coal (CFFC)	2	6	3	36.7
1.2.5	Conoco Zinc Halide Hydrocracking	2	4	3	30.0
1.3.1	Exxon Donor Solvent (EDS)	3	6	5	46.7
1.3.2	ADL Extractive Coking	2	8	3	43.3
1.3.3	Consol Synthetic Fuel (CSF)	5	5	6	53.3
2.1.1	Char-Oil-Energy-Development (COED), RMP	5	10	6	70.0
2.1.2	Char-Oil-Energy-Development (COED), FMC	5	10	7	73.3
2.1.3	Occidental Research Corp. Flash Pyrolysis	3	10	5	60.0
2.1.4	TOSCOAL	5	10	7	73.3
2.1.5	Lurgi-Ruhrgas	10	10	9	96.7
2.2.1	U.S. Steel Clean Coke	3	5	4	40.0
2.2.2	Coalcon	5	6	5	53.3
2.3.1	BNL Rotating Fluidized Bed	1	2	1	13.3
2.3.2	Short Residence Time (SRT) Coal Hydro-pyrolysis	2	6	3	36.7
2.3.3	Intermediate Coal Hydrogenation	2	5	3	33.3
2.3.4	Schroeder's Rapid Hydrogenation	2	6	3	36.7
2.3.5	BNL Flash Hydropyrolysis	2	5	2	30.0
2.3.6	Rockwell International Direct Coal Hydrogenation	2	6	2	33.3
3.1.1	Flame Sprayed Catalyst (FSC)	2	6	7	50.0
3.1.2	ARGE (Arbeit Germeinschaft, Ruhrchemie/Lurgi)	10	10	10	100.0
3.1.3	SYNTHOIL	10	10	10	100.0
3.2.1	High Pressure Methanol Process	10	10	10	100.0
3.2.2	ICI LP/LT Methanol	10	10	10	100.0
3.2.3	Lurgi LP/LT Methanol	10	10	10	100.0
3.2.4	Three-Phase Methanol Synthesis	2	10	3	50.0
3.3.1	M-Gasoline (Mobil)	2	10	3	50.0
4.1.1	Supercritical Gas Extraction	2	6	3	36.7
^a Weighting Factor: 3.33					

Table A-6 - Commercial Potential Rating

Process Number	Process	Criteria												Score (Maximum = 100)
		Complexity P1 0.855 ^a	Severity P2 0.945	Industrial Capacity P3 0.45	Feed Versatility P4 0.675	Product Versatility P5 0.675	Liquefaction Efficiency P6 0.90	Sulfur Reduction Efficiency P7 1.575	Conversion Efficiency P8 0.54	Product Premium P9 1.26	Product Compatibility P10 0.18	Product Hazard P11 0.945	Uncertainty Factor P12 1.0	
1.1.1	Solvent Refined Coal (SRC) I	4	3	10.0	7.5	5	7.9	8.5	1.8	5.6	0.8	2.0	5.3	55.1
1.1.1	Solvent Refined Coal (SRC) II	4	3	4.1	10.0	5	7.7	9.3	4.6	5.6	2.4	2.0	5.3	56.5
1.1.2	UOP Extraction Process	4	5	0.7	10.0	5	7.8	8.3	1.7	5.6	1.6	2.0	3.7	52.5
1.2.1	Gulf Catalytic Coal Liquids	6	3	5.9	10.0	6	6.8	9.2	7.3	7.8	5.6	2.0	4.3	63.0
1.2.2	SYNTHOIL	5	3	10.0	10.0	6	7.4	9.3	7.0	7.7	5.2	2.0	4.3	64.3
1.2.3	H-Coal	3	3	4.1	10.0	6	7.2	9.7	8.4	8.0	7.1	2.0	5.0	62.6
1.2.4	Clean Fuel from Coal (CFFC)	3	3	4.1	10.0	6	5.8	9.4	4.8	7.0	3.1	2.0	3.7	55.6
1.2.5	Conoco Zinc Halide Hydrocracking	2	3	4.5	10.0	6	7.7	9.7	5.8	7.1	5.7	2.0	3.0	57.6
1.3.1	Exxon Donor Solvent (EDS)	5	3	5.9	10.0	5	6.1	9.3	4.6	5.6	3.3	2.0	4.7	56.8
1.3.2	ADL Extractive Coking	6	7	4.1	10.0	6	5.3	2.5	10.0	5.6	10.0	2.0	4.3	53.5
1.3.3	Consol Synthetic Fuel (CSF)	2	5	4.8	10.0	6	5.7	9.6	9.0	8.3	8.1	2.0	5.3	63.6
2.1.1	Char-Oil-Energy-Development (COED), RMP	3	4	4.3	6.0	4	2.6	5.7	5.6	5.9	4.2	2.0	7.0	46.5
2.1.2	Char-Oil-Energy-Development (COED), FMC	3	4	4.1	6.0	4	2.5	3.7	5.0	5.8	3.6	2.0	7.3	42.9
2.1.3	Occidental Research Flash Pyrolysis	5	4	10.0	6.0	4	3.7	3.3	5.0	5.8	3.4	2.0	6.0	46.3
2.1.4	TOSCOAL	7	7	1.4	6.0	4	1.8	6.4	6.7	5.8	4.4	2.0	7.3	52.6
2.1.5	Lurgi-Ruhrgas	6	4	1.7	6.0	4	1.9	5.0	10.0	6.1	10.0	2.0	9.7	52.5
2.2.1	U.S. Steel Clean Coke	2	4	1.7	7.5	4	1.2	9.1	5.2	7.7	2.4	2.0	4.0	48.3
2.2.2	Coalcon	4	4	10.0	5.0	4	3.0	9.0	10.0	5.9	5.0	2.0	5.3	55.6
2.3.1	BNL Rotating Fluidized Bed	5	1	10.0	5.0	3	6.0	8.7	10.0	8.0	10.0	3.8	1.3	56.4
2.3.2	Short Residence Time (SRT) Coal Hydropyrolysis	5	1	10.0	5.0	1	1.6	9.9	10.0	8.0	10.0	4.0	3.7	55.6
2.3.3	Intermediate Coal Hydrogenation	3	1	10.0	5.0	3	5.2	8.7	10.0	6.6	8.1	2.6	3.3	57.7
2.3.4	Schroeder's Rapid Hydrogenation	4	3	1.7	5.0	3	6.0	8.7	10.0	7.2	10.0	3.1	3.7	54.4
2.3.5	BNL Flash Hydropyrolysis	5	1	0.0	5.0	3	1.6	9.9	10.0	7.3	10.0	3.2	3.0	50.1
2.3.6	Rockwell International Direct Coal Hydrogenation	5	1	10.0	5.0	3	4.4	7.6	6.1	5.6	0.0	2.0	3.3	46.6
3.1.1	F-T Flame Sprayed Catalyst (FSC)	6	10	0.7	10.0	5	4.0	10.0	10.0	6.9	10.0	9.6	5.0	74.3
3.1.2	F-T ARGE (Arbeit Germeinschaft, Ruhrchemie/Lurgi)	5	10	0.0	10.0	6	4.4	10.0	10.0	7.3	10.0	8.0	10.0	78.2
3.1.3	F-T SYNTHOIL	5	10	1.4	10.0	6	3.4	10.0	10.0	6.7	10.0	6.3	10.0	75.6
3.2.1	High Pressure Methanol Process	6	1	10.0	10.0	1	6.0	10.0	10.0	7.1	0.0	4.0	10.0	67.3
3.2.2	ICI LP/LT Methanol	6	6	2.4	10.0	1	6.0	10.0	10.0	7.1	0.0	4.0	10.0	68.6
3.2.3	Lurgi LP/LT Methanol	6	6	4.5	10.0	1	6.0	10.0	10.0	7.1	0.0	4.0	10.0	69.5
3.2.4	Three-Phase Methanol Synthesis	5	6	2.4	10.0	1	6.0	10.0	10.0	7.1	0.0	4.0	5.0	62.7
3.3.1	M-Gasoline (Mobil)	3	7	0.3	10.0	1	5.3	10.0	10.0	10.0	10.0	6.0	5.0	67.7
4.1.1	Supercritical Gas Extraction (SGE)	7	3	2.4	10.0	1	3.7	4.2	0.0	3.7	0.0	2.0	3.7	37.5

^aWeighting Factor

Figure A-1 - Industrial Capacity Rating (P3)

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 plant 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 Orleans, 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 stipula-

tion 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 transporting 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	1.03	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

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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:

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

Condensed water is treated and used for cooling tower make-up and coal slurring. Hydrocarbons produced from the reaction are separated in the gas plant to produce SNG, a C₃-C₅ liquids stream as depentanizer overhead, and C₆-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 synthesized C₆-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 solution 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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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	<u>5.2</u>
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	15,790	106,572	551
Sulfur	1,759		14
Na ₂ SO ₄	<u>25</u>	<u> </u>	<u> </u>
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	<u>0.27</u>
Total Constructed Cost	2.14
Engineering, Home Office, Fee and Sales Tax at 125%	<u>0.26</u>
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	<u>0.04</u>	<u>0.04</u>	<u>0.04</u>
Total Depreciable Investment	2.23	2.20	2.17
Working Capital	<u>0.17</u>	<u>0.18</u>	<u>0.13</u>
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	0.03
Maintenance Materials and Contract Labor	0.19	0.17	0.30
Catalysts and Chemicals	<u>0.03</u>	<u>0.03</u>	<u>0.04</u>
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	<u>0.14</u>	<u>0.13</u>	<u>0.22</u>
Total Labor Costs	0.36	0.34	0.57
G and A Overhead	0.03	0.03	0.04
Property Taxes and Insurance	<u>0.22</u>	<u>0.20</u>	<u>0.36</u>
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	0.55
Operations	0.57	0.54	0.72
By-product Credit	<u>-0.06</u>	<u>-0.06</u>	<u>-0.06</u>
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	<u>1.30</u>	<u>1.28</u>	<u>1.32</u>
Product Total	1.41	1.39	1.43
By-products			
Sulfur	0.08	0.07	0.07
Sodium Sulfate	<u>0.01</u>	<u>0.10</u>	<u>0.10</u>
By-product Total	0.09	0.08	0.08
Total Annual Revenue Ratio	<u>1.50</u>	<u>1.47</u>	<u>1.51</u>
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.

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

THERMAL EFFICIENCY: $600 \times 100 = 68\%$
880

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THE RALPH M. PARSONS COMPANY PASADENA, CALIFORNIA				JOB NO. 5658-1	DWS. NO. R-06-FS-1			REV. 0

	TONS/DAY	MMSCFD	MMMBTU/D (HHV)
FEED COAL	35,980	—	900
HYDROGEN	6,390	1,716	600
SULFUR	1,189	—	—
TOTAL	7,579	—	600

THERMAL EFFICIENCY: $\frac{600}{900} \times 100 = 67\%$

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

Table D-1 - Ratios of Total Capital Investment
SNG and H₂ 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	<u>0.03</u>	<u>0.03</u>	<u>0.02</u>
Total Depreciated Investment	1.56	1.59	1.34
Working Capital	<u>0.12</u>	<u>0.12</u>	<u>0.10</u>
Ratio to Total Capital Investment to SRC-II	1.68	1.71	1.44

Table D-2 - Sensitivity 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	<u>2.1</u>	<u>3.6</u>	<u>1.0</u>
Total Material, Supplies, and Utilities	13.5	15.2	11.7
Labor			
Operating Labor and Supervision	1.7	1.3	1.4
Maintenance Labor and Supervision	5.1	5.3	4.7
Payroll Burden	2.4	2.2	2.4
Plant Overhead/Administration	<u>5.5</u>	<u>5.4</u>	<u>4.7</u>
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	<u>100.0</u>	<u>100.0</u>	<u>100.0</u>
Operating Costs Ratio to SRC II	1.29	1.27	1.20

Table D-4 - Ratios of Unit Btu RPSP
SNG and H₂ 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	<u>-0.04</u>	<u>-0.04</u>	<u>-0.04</u>
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
KEY ECONOMIC COST DATA

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	-	-