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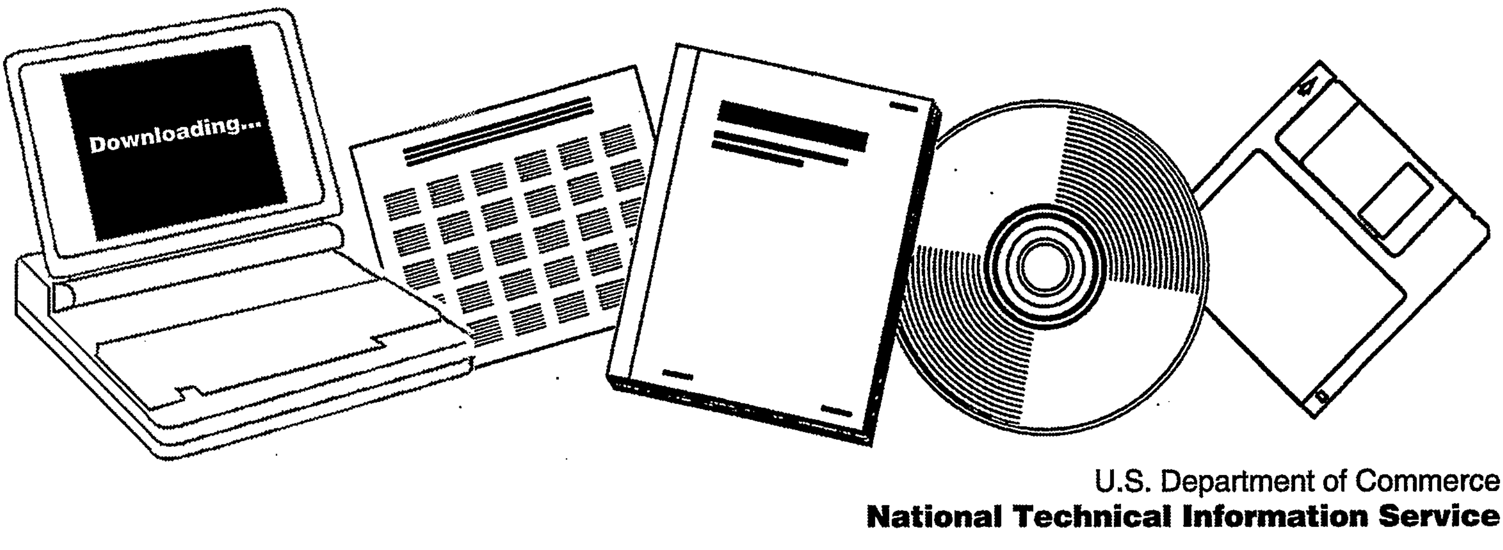

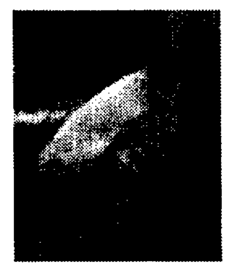

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PROJECT POGO
TOTAL COAL UTILIZATION
COG REFINERY DESIGN CRITERIA

R & D REPORT NO. 114, INTERIM REPORT NO. 5

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ABBREVIATIONS

BPD	barrel per day
CF	cubic foot
COED	coal-oil energy development
COG	coal-oil-gas
CSF	consolidation synthetic fuel
D	day
DCF	discounted cash flow
DDB	double-declining balance
EP	end point
FCI	fixed capital investment
F/T	Fischer-Tropsch
H-Coal	hydrogenated coal process

ABBREVIATIONS (Contd)

HHV	higher heating value
LPG	liquified petroleum gas
LTSD	long ton per stream day
MAF	moisture- and ash-free (coal)
M	thousand
MF	moisture-free (coal)
MM	million
MMM	billion
O/G	Oil/Gas
POGO	power-oil-gas-other
psi	pound per square inch
psig	pound per square inch gauge
PU	pyrolysis unit
ROI	return on investment
ROM	run-of-mine (coal)
RPSP	required product selling price
S	sulphur
SCF	standard cubic foot
SCFD	standard cubic foot per day
SD	stream day
SNG	substitute natural gas
SRC	solvent refined coal
syngas	synthesis gas
TPD	ton per day
TPSD	ton per stream day
wt %	weight percentage

SECTION 1

INTRODUCTION

The objective of this work is to develop design criteria for a conceptual design/economic evaluation for a multiproduct complex to convert coal to electric power, oil, gas, and other products. ERDA has designated this multiproduct complex, POGO, an acronym for power-oil-gas-other. POGO is an outgrowth of earlier work on a design concept referred to as coal-oil-gas (COG). Therefore, the POGO concept uses multiple processes in a preferred combination to produce a broad spectrum of environmentally acceptable fuels, plus other products, that will be economically competitive with alternative sources of these products.

The objective was achieved by analyzing the capabilities of the major generic types of liquefaction processes and then by comparing the projected technical and economic performances. The next step was to compare the predicted performance of a number of potentially viable candidate combinations of processes and to recommend the preferred process combination, plus preliminary design criteria. The intent is that the resulting complex would use the best available coal conversion processes in combinations such that the byproducts or wastes of one process would form inexpensive raw materials for another process. In this manner, what might have been expenses could be turned into savings, and the final product cost could be lower than that possible with a single process plant.

The program was designed to review and analyze at least one candidate process in each of the following generic liquefaction categories:

Category	Process Reviewed
Hydroliquefaction	
Noncatalytic	SRC I
Pseudocatalytic	SRC II, Oil/Gas
Catalytic	H-Coal, Synthoil
Donor Solvent	CSF
Pyrolysis	
Direct	COED
Hydropyrolysis	Coalcon
Indirect	Fischer-Tropsch

SECTION 2

SUMMARY

Design criteria have been developed for a conceptual coal conversion complex to produce environmentally clean liquid and gaseous fuels, plus electrical power. A summary is presented of this design criteria development program.

The objectives for the program are best described by reference to the work statement for the POGO conceptual design/economic analysis assignment that specified three phases for the work:

(1) The contractor shall perform and submit preliminary analyses of existing processes and make recommendations from which the Government shall select the better combinations.

(2) Complete conceptual design of the processes selected under phase 1.

(3) Optimize concept design.

The program plan included design development for processing three coals in three different geographical areas of the United States. The intent is to study preferred process configurations and to optimize the results.

This report summarizes the results of the work that fulfilled the obligations under phase 1, described above. The design criteria presented here are now being used for development of the phase 2 conceptual design.

The development of the design criteria required analyses of candidates from all major generic types of coal conversion technologies, plus a number of potentially viable combinations of processes. The following factors should be noted when using the results reported:

- The resources available for the analyses were limited, both in personnel manhours and in cost, with regard to the broad scope of the objectives.
- All technologies that were reviewed are still under development.
- The information available represents the status at a point in time that is based on information available to an investigative team.
- The data and information came from many sources, and comparisons must rationalize the nature and quality of the input; i.e., some of the predicted economics are based on completed comprehensive conceptual design/economic evaluations by independent designers, while others are based on information supplied by process developers.

The results presented here, placed in their proper perspective based on the method of development, provide a broad display of major characteristics of coal liquefaction processes now under development and their potential, "relative" economics. Significance should be attached to the reference to relative values rather than absolute values, which must come from more detailed assessment efforts. The results provided the basis for a systematic and complete program designed to select a preferred process configuration for a subsequent, detailed conceptual design assignment. Key elements of the program are summarized in the following paragraphs.

The design criteria program consisted of the following logic pattern:

- Preliminary screening of existing processes in which approximately 85 combinations and permutations of individual coal conversion processes, plus supporting facilities, were considered using semiquantitative screening procedures.

- A preliminary review/analysis of nine coal conversion complexes included at least one candidate process from each of the major generic coal liquefaction categories.

  - Process descriptions, block flow diagrams, heat and material balances, and preliminary economics were developed for each candidate process.

  - Process and economic analysis results developed for prior conceptual designs, plus two in-progress designs, were used as input to this phase of the program.

  - Results: The solvent refined coal (SRC) type of hydroliquefaction processes showed promise as a low-cost, clean fuels producer. The consolidation synthetic fuel (CSF) type of donor-solvent process appeared to be of a slightly higher cost. Low-pressure pyrolysis appeared to be a high fuel cost route; other candidate processes arrayed themselves in intermediate-projected product-fuel cost positions.

- A short list of four high potential process/process combination candidates was developed after analysis of the results of the preliminary review/analysis program, plus further analytical work. The list consisted of:

Case	Process Configuration
I	CSF with low-temperature carbonization
II	Flash pyrolysis, CSF, and Fischer-Tropsch
III	Flash pyrolysis and Fischer-Tropsch
IV	Flash pyrolysis and SRC

- The flash pyrolysis was included to "skim" easily recoverable high Btu gas and tar from the feed coal by a pressurized flash pyrolysis step and to produce a char that, in turn, is gasified to produce the necessary syngas and/or hydrogen for use in liquefaction. The flash pyrolysis also permitted the development of a method to exclude the troublesome filtration step from SRC processing. The elimination of this step is important because filtration of the fine (1 to 10 micron particle size), unreacted coal-plus-ash solids is expensive and a difficult, commercial operation. This point is discussed in OCR R&D Report No. 82, Interim Report No. 1 (Ref. 1).

- Preliminary design configurations and economics were developed for each of the four cases listed above, plus a suggested second-generation U.S. Fischer-Tropsch plant and Oil/Gas (an SRC II-based process) designs that were in progress at Parsons. The technical and economic results were analyzed.

• A preferred process configuration was selected that had been based on the results of the programs' steps described above. Case IV was selected as the recommended configuration; Figure 4-4 of Section 4 shows the block flow diagram for this case.

• A design criteria document was developed for Case IV and is presented in this report. It is intended to:

  - Describe key elements of the design that will permit users to anticipate size, product state, and general characteristics of the resulting facility.

  - Permit designers to proceed with their objectives and work.

The completion of the selection process has provided the basis for proceeding with the development of the conceptual design/economic evaluation of the POGO complex. Preliminary economic analyses were based on information that was available to the investigators in mid-1976. Limited schedule and budget were available to meet the broad scope of the objectives, which included the analysis of all potentially viable coal conversion candidate process combinations, plus advanced electrical power generation facilities. Within these constraints, a systematic program of analysis, a significant amount of technical and economic analysis supporting the assembled information, and a decision regarding the preferred design criteria have been completed.

Information presented in this report should be used with full recognition of the manner and purpose of its development. Comparisons were based on a number of technologies that were only under development and on information available to a particular investigative team.

Parsons recommends that assessment of candidate processes be continued and expanded as more information becomes available. Future emphasis should be placed on product characteristics/marketability, process/thermal efficiency comparison between alternatives, and materials of construction/equipment performance; however, other factors must also be considered. It is suggested that the type of preliminary assessment presented here be extended and that increasingly sophisticated analysis procedures be applied as the quantity and quality of informational input from the development programs increase.

SECTION 3

PRELIMINARY REVIEW/ANALYSIS

INTRODUCTION

The initial effort assembled, reviewed, and analyzed information for each of nine candidate liquefaction processes, exclusive of the Fischer-Tropsch and Oil/Gas designs, which were being developed independently, at the same time, by Parsons. The following information was developed:

- Date sources and status
- Process description
- Preliminary technical analysis
- Preliminary block flow diagram
- Heat and material balances
- Preliminary economic analysis

This information became factors for the following eight processes:

- SRC I using hydrogen as the hydroliquefaction agent
- SRC I using hydrogen plus carbon monoxide (syngas) as the hydroliquefaction agent
- SRC II (slurry recycle) using hydrogen
- SRC II using syngas
- H-Coal
- Synthoil
- CSF
- COED

Other input to the review/analysis program consisted of equivalent information for the Coalcon hydrocarbonization process, which was developed by Coalcon (Ref. 5) and Dravo (Ref. 6) under an ERDA sponsorship, and the Parsons Fischer-Tropsch and Oil/Gas information available at that time from in-process design work.

Every attempt was made to recognize and define the differences in the nature of the information used. Designs listed in Refs. 1, 2, 3, and 4 used much more of Parsons effort and manhours than those used for the H-Coal, Synthoil, CSF, and Coalcon processes. An attempt was made to assign proper weight to the efforts of other organizations on these processes to develop the published data available. The Coalcon design information, other than reduction of plant capacity to conform with the objectives of this review, was used as published and therefore differed in some respects from the others. These differences in data sources between the processes must be recognized when making comparisons and drawing conclusions in the limited effort analysis program reported here.

Procedures to speed the development of the process comparison information included:

(1) A computer model of a "standard" coal conversion complex to calculate component balances, estimates of utility balances, and fixed capital investments (FCI) for units of the complex that are based on unit capacities and information from completed designs (Refs. 1 and 2).

(2) Computer-assisted FCIs where appropriate.

(3) Computer-assisted profitability analyses.

To screen the POGO process combination candidate, a very preliminary, semi-quantitative analysis of approximately 85 combinations and permutations of the generic process classifications, plus feedstock preparation and downstream processing, was completed during the early days of the review. This was pursued as far as practical within the limits of time and manhours available for the analysis. The results were used as guidance for subsequent analyses efforts to be described. Further details regarding the bases used for the comparison are presented in the following paragraphs.

Process Comparison Basis

(1) Feed coal = Illinois No. 6, West Virginia (Ireland), or Kentucky, with analyses as follows:

Proximate Analysis	Illinois No. 6	W. Virginia	Kentucky
Moisture, wt % wet basis	2.7	2.7	2.7
Ash, wet % dry basis	7.3	7.3	7.3
Volatile matter, wt % MAF basis	46.4	45.0	45.5
Fixed carbon, wt % MAF basis	53.6	55.0	54.5
Higher heating value, Btu/lb	12,536	12,640	12,536
Ultimate Analysis, Wt % MAF Basis			
Carbon	78.6	79.5	79.1
Hydrogen	5.4	5.6	6.0
Nitrogen	1.5	1.5	1.6
Sulfur	4.3	4.9	5.7
Oxygen	10.2	8.5	7.6

(2) Feed coal preparation and processing downstream of the prime coal conversion step used basic flow plans similar to Ref. 1 wherever possible. Characteristics were:

  (a) Plant capacity was 25,000 tons per stream day (TPSD) of feed coal to the prime coal conversion reactor.

  (b) Dissolver sections removed ash and unconverted coal by filtration; the filter cake was dried.

  (c) Hydrogen was produced in an entrained, slagging, 2-stage gasifier operating at approximately 200 pounds per square inch gauge (psig).

  (d) Sweet shift reactors were used.

  (e) Pipeline quality gas was produced by methanation of the final product.

  (f) Where proper yield data was available, the prime fuel products were pipeline gas (SNG), LPG, naphtha, and fuel oil.

  Chemical byproducts such as ammonia, phenol, and cresols were not considered for this study because reliable yield data was generally not available. The fact that these materials were excluded here does not preclude their contribution to production quantities and economics in future, more detailed, analyses.

(3) The plant produced all captive fuel and power requirements in an air-blown, low-pressure, 2-stage gasifier. Raw water was supplied from a nearby river.

(4) Utilities and FCI for individual units were based on Ref. 1 estimates escalated to fourth quarter 1975 costs; this coincides with the time frame of the in-progress Fischer-Tropsch and Oil/Gas economic analyses.

(5) SNG was produced at 1,000 psig, 1,000 Btu/SCF HHV, 2% H_2 max. and 0.1% CO max.

(6) Operating manpower was estimated from Ref. 1 results with an adjustment for plant capacity and complexity.

(7) A cost of run-of-mine (ROM) coal of $8.75 per ton was used. This was representative of transfer prices from large, conceptual, captive strip mines being developed at that time. This transfer price includes coal mine return on investment (ROI). In general, the sensitivity of the required product selling price (RPSP) to coal cost at 12% discounted cash flow (DCF) rate of return is about 0.3. This means that if the cost of coal doubled to ROM $17.50 per ton, the RPSP would increase by about 30%.

Exceptions to these general bases were:

- COED - principal products are syncrude and SNG.
- Fischer-Tropsch - products are SNG, LPG, naphtha, diesel fuel, and fuel oil.
- CSF - feed is Pittsburgh No. 8 seam coal.

Economic Comparison Basis

The first-order economic analysis consisted of the development of very preliminary estimates for the required FCI, total capital requirements, operating costs, and the RPSP to provide a 12% DCF rate of return for a specified project structure. Details of the basis used for the preliminary economic analysis work are shown in Table 3-1, at the end of this section.

Candidate Processes

Preliminary technical and economic analyses were completed for the following nine processes during this phase of the campaign:

SRC - I using approximately 94% purity hydrogen as hydroliquefaction agent

SRC - II using syngas (45% hydrogen and 50% carbon monoxide) as hydroliquefaction agent

SRC - II using hydrogen

SRC - II using syngas

H-Coal

Synthoil

CSF - donor solvent process

COED - to produce syncrude and SNG

Coalcon

Information from these process/economic analyses was supplemented by the results of the Fischer-Tropsch and Oil/Gas designs as they became available.

The results for the nine processes are presented in Appendix A, which summarizes for each process:

- Data source and status
- Process description

* Block flow diagram

* Key heat and material balance factors

The results of the review/analysis are summarized in the subsequent paragraphs.

Plant Capacities/Energy Balances

The total quantity of coal feed to each plant is summarized in Table 3-2. Also presented is the indicated thermal efficiency.

The projected thermal efficiencies for the SRC-type processes are indicated to be in the range of 73 to 76%, while those for the H-Coal and Synthoil are presented as 64 to 66%. The initial conceptual design published for an SRC design in this series (Ref. 1) was also projected to have about a 65% thermal efficiency. Subsequent design and analysis work over the past 3 years has defined ways that the efficiency might be improved to the 75% range. The liklihood exists that a similar effort for the catalytic hydroliquefaction processes might also increase the efficiency to the same range. The detailed analyses for these processes was beyond the scope of this work and the information available to us; the efficiencies listed in Table 3-2 are therefore based on information published by the process developers.

PRODUCT CHARACTERISTICS

A general summary of projected product characteristics for the candidate processes is shown in Table 3-3. The differences must be recognized in making comparisons of the candidate processes.

PRELIMINARY ECONOMIC COMPARISON

The relative FCI and the RPSP are shown in Table 3-4. The FCIs for all processes except Coalcon were developed from adjusted investments of units that had been estimated in detail for Refs. 1 and 2; the Ref. 3 Coalcon estimate was scaled down from approximately 50,000 TPSD of coal feed to 25,000 TPSD, based on an analysis of the appropriate scaling factors for the separate units contained in the design. The RPSP (cost of fuel) for all processes was estimated by using the economic parameters listed in Table 3-1.

The results of this preliminary assessment indicated that, using information available at that time, the technologies array themselves in the following order of increasing RPSP, expressed as dollars per million Btu ($/MM Btu); as noted, this analysis does not recognize probable product market values caused by differences in product characteristics.

* SRC II - H_2 gas

* SRC I - H_2 gas

* SRC II - Syngas

* SRC I - Syngas

- CSF
- H-Coal
- Synthoil
- Coalcon
- COED

For this type of preliminary analysis, the projected relative costs have much more significance than a projected absolute RPSP. However, for general guidance purposes, an approximate required revenue and an RPSP based on information available at that time were generated for the SRC II - H_2 gas process and is shown in Table 3-5. Again, this represents the results of a very preliminary assessment at that point in time; it has been subsequently revised as more information/effort was applied.

SENSITIVITY FACTORS

Preliminary judgments regarding the sensitivity of the preliminary economics to process and produce characteristic factors are discussed here.

Preliminary analysis indicates that, in general, capital associated costs contribute 60 to 75% of the RPSP, coal contributes 25 to 35%, and other operational costs, 4 to 7%.

Effect of Capital Costs

For the processes considered, the variation in the FCI is the principal contributor to the cost of fuel produced; it is also the major contributor to variations in the cost between processes. Sensitivity of the RPSR to variations in the FCI is about 0.8; i.e., a 10% decrease in the FCI will result in about an 8% decrease in the RPSP.

The projected high FCIs for the catalytic processes are due primarily to their high hydrogen consumption, which, in turn, requires large gasification facilities.

Effect of Plant Efficiency

The effect of the thermal efficiency of the process is reflected in the cost contribution of the coal feedstock. Sensitivity of the RPSP to the efficiency of coal conversion is approximately one-half the FCI sensitivity.

The projected thermal efficiencies of approximately 75% of the SRC processes are the result of significant analysis and process improvement effort (Ref. 8); a 1973 design projected about 65% efficiency with the basis for this efficiency explained (Ref. 1). The efficiencies used for the catalytic hydroliquefaction processes in this study originated from published information by the process developers. The probability exists that their thermal efficiencies could be improved by further design analysis efforts.

Effect of Product Type

The average product cost rises as more gas and/or more light liquids are produced. An application of the SRC I processes is to produce primarily a high melting point, dissolved-coal product. The SRC II processes can be used to produce primarily a liquid product that is usable as liquid fuel. The combined liquid products provided by the catalytic process could be called a synthetic crude oil. The economic analysis presented does not attempt to compensate for these differences.

CONCLUSIONS

At present, the noncatalytic SRC process is indicated to be a low-cost producer of fuel products. The catalytic processes can produce lower sulfur and lighter products but at a higher cost. Lower sulfur products could also be achieved by further hydrotreating the SRC products. In general, gas and light hydrocarbons have higher production costs, on a heating value basis, than heavy liquids.

This review/analysis presents the characteristics and projected potential of the processes at a point in time when each is still under development; the review was completed with a limited manhour effort, considering the broad scope of the assessment. The results are considered as a contribution to the task objective of selecting a preferred process configuration for the POGO design, based on information available to Parsons.

Table 3-1 - Economic Assessment Basis, Preliminary Process Comparison

Basis	Fourth Quarter, 1975
Project life	20 years of plant operation
Construction period	4 years
Operating factor	330 stream days/year
Startup expenses	2% of capital investment
Project financing	100% equity

Working capital is based on the following factors:

- 30-day feed coal inventory
- 30-day inventory of finished product
- Cost of spare parts inventory is equal to 4% of major equipment cost
- 30-day accounts receivable
- 30-day budget for current expenses
- 30-day credit for accounts payable

Annual Operating Costs	
Coal feedstock	\$8.75/ton, ROM
Water	\$0.10/M gal
Operating labor wages	\$6.50/h, \$13,000/man-year
Operating labor supervision	15% of operating labor
Annual maintenance cost	4% of FCI
Annual burden cost	35% of labor
Plant overhead	60% of operating labor (including payroll burden)
Annual G and A overhead allowance	1.5% of manufacturing cost
Annual cost, property tax, and insurance	2.75% of FCI
Depreciation	13-year period, double-declining balance (DDB) method
Federal income tax	48% of profit before tax
State income tax	4% of profit before tax
Investment tax credit	10% of FCI
Discounted cash flow rate of return	12% after tax

Table 3-2 - Summary, Energy Balances

Process	Coal Feed TPSD		Thermal Efficiency (%)
	To Prime Converter	Total	
SRC I - H_2 gas	25,000	36,200	76
SRC I - Syngas	25,000	36,900	74
SRC II - H_2 gas	25,000	39,340	74
SRC II - Syngas	25,000	40,550	73
H-Coal	25,000	50,920	66
Synthoil	25,000	46,200	64
CSF	25,000	28,350	77
COED	25,000	25,000	55
Coalcon	25,000	25,000	62

Table 3-3 - Summary, Projected Product Characteristics

Process	Product Characteristics				
	SNG	LPG	Naphtha	Diesel Fuel	Fuel Oil
SRC I - H_2 gas	Pipeline quality	Mixed	400°F EP 1 ppm sulfur	-	0.8 wt% sulfur 400°F+
SRC I - Syngas	Pipeline quality	Mixed	400°F EP 1 ppm sulfur	-	0.8 wt% sulfur 400°F+
SRC II - H_2 gas	Pipeline quality	Mixed	400°F EP 1 ppm sulfur	-	0.5 wt% sulfur 400°F+
SRC II - Syngas	Pipeline quality	Mixed	400°F EP 1 ppm sulfur	-	0.5 wt% sulfur 400°F+
H-Coal	Pipeline quality	Mixed	400°F EP 1 ppm sulfur	-	0.7 wt% sulfur 400°F+
Synthoil	Pipeline quality	-	400°F EP 1 ppm sulfur	-	0.3 wt% sulfur 400°F+
CSF	Pipeline quality	-	400°F EP	-	0.3 wt% sulfur 400°F+
COED	300 Btu/CF 250 psig	-	-	-	0.1 wt% sulfur
Coalcon	Pipeline quality	Mixed	(light oil) 2.2 wt% sulfur	-	Not reported
Fischer-Tropsch	Pipeline quality	Butane	380°F EP Nil sulfur and N_2	Nil sulfur and N_2	Nil sulfur
Oil/Gas	Pipeline quality	Propane and Butane	400°F EP 1 ppm sulfur 5 ppm nitrogen	-	400°F+ 0.4 wt% sulfur

Table 3-4 - POGO Plant Process Comparison Summary, Yield and Cost (Coal Source: Illinois)

Yields									
Feed coal to prime coal conversion step, TPD	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000
Total feed coal, TPD (2.7% H_2O)	39,340	36,200	40,550	36,900	28,350	50,920	46,200	25,000	25,000
Coal HHV Btu/lb 2.7% H_2O	12,536	12,536	12,536	12,536	12,536	12,536	12,536	-	12,536
Basic process	SRC II H_2 gas	SRC I H_2 gas	SRC II Syngas	SRC I Syngas	CSF	H-Coal	Synthoil	Coalcon	COED
Type of gasifier	Entrained	Entrained	Entrained	Entrained	Entrained	Entrained	Entrained	Fluidized	Fluidized
Products (percentage of coal HHV)									
Gas	18.2	13.2	19.6	14.0	12.7	17.2	21.6	26.2	26.2
LPG	1.0	2.2	0.5	1.1	–	3.9	–	4.1	-
Naphtha	3.0	4.4	2.9	4.3	10.6	12.5	8.6	9.5	-
Fuel oil	50.5	54.9	49.0	53.8	52.8	31.2	32.8	21.3	27.9
Sulfur	0.9	0.9	0.9	0.9	1.2	0.9	0.9	1.0	0.8
Total	73.6	76.6	72.9	74.1	77.3	65.7	63.9	62.1	54.9
Sulfur in fuel oil (wt%)	0.5	0.8	0.5	0.8	0.3	0.7	0.3	Not Reported	0.1
Cost Relative to the Base Process of SRC II, H_2 Gas									
Relative contribution to product cost (%)									
Capital	59.8	60.9	60.5	61.3	65.9	61.3	69.5	63.4	74.0
Operations	5.4	5.5	5.2	5.5	6.0	6.9	7.1	6.9	3.5
Coal	37.3	36.0	36.8	35.6	31.6	34.2	32.7	33.5	24.8
Byproducts	-2.5	-2.4	-2.5	-2.4	-3.5	-2.4	-2.3	-3.8	-2.3
Relative fixed capital investment	1.00	0.97	1.06	1.01	0.94	1.46	1.41	0.75	1.17
Relative fixed capital investment per Btu product	1.00	1.03	1.04	1.06	1.25	1.26	1.38	1.44	2.33
Relative fixed capital investment per daily ton feed coal	1.00	1.06	1.03	1.08	1.31	1.13	1.20	1.18	1.84
Relative cost of fuel product, Btu basis	1.00	1.01	1.03	1.03	1.13	1.29	1.31	1.35	1.90

Table 3-5 - POGO Plant, Approximate Economic Factors
SRC II - H_2 Gas Process

Fixed Capital Investment	$1,100,000,000	
Economic Factor	Required Revenue	
	$MM/Yr	$/MM Btu
Capital-associated items	355.	1.50
Raw material	175.	0.75
Other operational costs	25.	0.10
Byproduct credit	-10.	-0.05
Total	545	2.30

SECTION 4

SHORT LIST SELECTION

The results of the preliminary review/analysis were used as a background for the selection of a short list of process configurations for further detailed analysis. The logic pattern was to select a preferred configuration based on the results of the short list analysis. The results of this phase of the program are described in this section.

The preliminary results available from in-progress conceptual designs for Oil/Gas and advanced Fischer-Tropsch processes were added to the nine processes listed in Table 3-4. The Oil/Gas is an SRC II-based process that produces significant SNG, while the Fischer-Tropsch design used flame-sprayed catalyst reactor systems for key reaction steps. Both of these conceptual designs incorporate some process steps and equipment items currently under development, and their commercialization depends on the successful completion of these developments. Descriptions of these two processes as they existed at the time of this study are presented in Appendix B. Further work continued on each of the two designs, and improvements were made prior to the finalization and publication of the ERDA R&D Reports (Refs. 7 and 8).

SHORT LIST

Approximately 12 candidate processes and process combinations were further screened prior to the selection of the short list. The process types showing better economics in Table 3-4 were emphasized. Configurations using multiple coal conversion process steps dominated the short list.

The following four process configurations were selected for the short list analysis program. In each case, electrical power generation facilities adequate to supply captive requirements, plus 1,000 MW for sale, were included.

Case	Process Configuration	Block Flow Diagram Figure No.
I	CSF with low temperature carbonization	4-1
II	Flash pyrolysis, CSF, and Fischer-Tropsch	4-2
III	Flash pyrolysis and Fischer-Tropsch	4-3
IV	Flash pyrolysis and SRC	4-4

The block flow diagrams are located at the end of this report section.

The reasons for selecting these four cases are:

- Case I: The CSF donor process has definable assets. The evaluation done during the preliminary review/analysis program previously described and the results presented in Table 3-4 included a limited amount of product refining. The objective of the POGO deisgn is to produce marketable products. The short list program, therefore, developed extensive revisions to the earlier configuration to achieve this objective. This work also served as a source of information for Case II.

- Case II: Addition of a pyrolysis process to the CSF scheme was intended to provide a moderate cost procedure to produce a high Btu gas, a tar, and a char. The purpose was to assess the potential technical and economic benefits that were to accrue from combining a donor solvent process with an operation that would inexpensively produce liquid and gaseous products from fresh coal and also produce char required for the generation of hydrogen.

The selection of an appropriate pyrolysis process was necessary. Potential choices included:

(1) Low-pressure, fluid-bed pyrolysis, such as the COED process.

(2) Near atmospheric pressure flash pyrolysis, such as the Occidental Research process.

(3) Fluid-bed hydropyrolysis.

(4) Several proposed, but undeveloped, entrained hydropyrolysis processes at an elevated pressure ($\simeq$1,000 psig) and temperature.

(5) Pressurized flash pyrolysis.

General comments about these processes include:

(1) Low-pressure, fluid-bed pyrolysis has been proven to be a high cost process. Furthermore, it produces a synthesis gas at low pressure, requiring costly and utility-intensive compression equipment. Therefore, it was not considered appropriate for POGO.

(2) Near atmospheric flash pyrolysis processes provide a potentially inexpensive means for pyrolysis, which has the further advantage of improving liquid product yield by means of a short contact time at pyrolysis temperature.

Operation near atmospheric pressure, however, requires either cooling and pressurizing of char or gas compression, both inherently expensive steps.

(3) Pressurized fluid-bed hydropyrolysis operates at an appropriate pressure for use in POGO. Operation of this system, using rather coarse coal, directs the gasifier to be a fluidized bed as well, if char

depressurizing, cooling, crushing, and repressurizing is to be avoided. The fluid-bed hydropyrolyzers and gasifiers are considered inherently more complicated and costly than dilute phase entrained-type reactors.

(4) A typical pressurized-entrained hydropyrolysis process was considered; it was concluded that the current developments are directed to severe hydrogenation, which does not leave adequate char for a plant hydrogen balance. It is suggested that the researchers/developers continue to define the kinetics to permit, where appropriate, the use of partial conversion to liquids while retaining enough unreacted char to serve as a hydrogen precursor.

The points just mentioned indicate that the two hydropyrolysis processes might be suitable for POGO, and flash pyrolysis would have advantages if it could be conducted at an elevated pressure. Operation of a flash pyrolysis at elevated pressure would reduce liquid yields (Ref. 9), but because flash pyrolysis characteristically produces a high liquid yield, it was decided to estimate the changes in yields with increased pressure and compare pressure flash pyrolysis with hydropyrolysis.

A brief engineering estimate was made of the equipment in the pyrolyzing sections only, and conversion costs for gas and liquid products were estimated, based on a 10% ROI after taxes. The results given in Table 4-1 show that Coalcon-type hydropyrolysis is approximately 15% higher in required product price than flash pyrolysis, and high pressure (about 1,000 psi) hydropyrolysis is 30% higher than flash pyrolysis. A cost difference of 15% was considered significant. Pressurized flash pyrolysis was therefore selected for use in POGO, based on the information available at the time.

- Case III: Flash pyrolysis and Fischer-Tropsch. The use of pressurized flash pyrolysis serves a purpose similar to that just described for Case I. Here, the char can be used to produce the snygas required for the Fischer-Tropsch reaction. The tar produced in pyrolysis supplements the liquid fuel production from the Fischer-Tropsch section. The plant can produce both aliphatic- and aromatic-based fuels.

- Case IV: Flash pyrolysis and SRC. The pressurized flash pyrolysis was combined with the SRC-type hydroliquefaction to provide efficiency in augmenting the fuels production from hydroliquefaction.

In each case, low Btu fuel gas was produced in a low-pressure, air-blown gasifier. It must be noted that in the final POGO design, the type of feed to the gasifier, pressure, and other design factors will be studied. The feeds to Cases I, II, and III are ROM coal that was beneficiated to produce high and low ash fractions. The low ash-high volatile fraction was fed to the pyrolysis, hydroliquefaction, and donor solvent conversion processes, while the high ash fraction was used in the low Btu gas generator. Conversely, the feed to Case IV was clean, sized, unfractionated coal. It is believed that use of the benefication process for Case IV would improve the projected economics.

The material balances, plus relative FCIs and projected RPSP for the Oil/Gas, Fischer-Tropsch, and Cases I-IV are presented in Table 4-2. Observations of the tabulated results include:

- The projected RPSP for Oil/Gas and Cases I-IV varies over a narrow range; the difference is about 10%.

- Case I (CSF) with a 75% thermal efficiency is indicated to require a slightly higher product selling price than the SRC-based Oil/Gas plant; this is the same relationship indicated by the preliminary review/analysis program results.

- Pyrolysis plus SRC is equal to or of a slightly lower cost than SRC alone (Case IV vs. Oil/Gas plant).

- Fischer-Tropsch plus pyrolysis requires a lower product selling price than Fischer-Tropsch alone (Case III vs. Fischer-Tropsch plant).

- CSF plus pyrolysis requires a slightly higher product price than CSF alone (Case I vs. Case II). This comparison is confounded by the inclusion of Fischer-Tropsch in Case II. However, Fischer-Tropsch is a small part of this plant, and it is felt that the indicated result is a true observation, caused by the fact that, according to published information, CSF itself is essentially self-sufficient in char to produce the necessary hydrogen, while SRC requires more hydrogen than its residual char can produce. Thus, CSF plus pyrolysis appears to offer no potential economic incentives.

- SRC plus pyrolysis (Case IV) indicates the lowest RPSP of the four short list cases.

  - The use of beneficiated coal feed is expected to further improve the projected economics.

  - Available information indicates that the thermal efficiency can be improved from the indicated 71% level, which should be compared with the Oil/Gas case.

The results of this analysis were used to develop a recommended configuration for POGO.

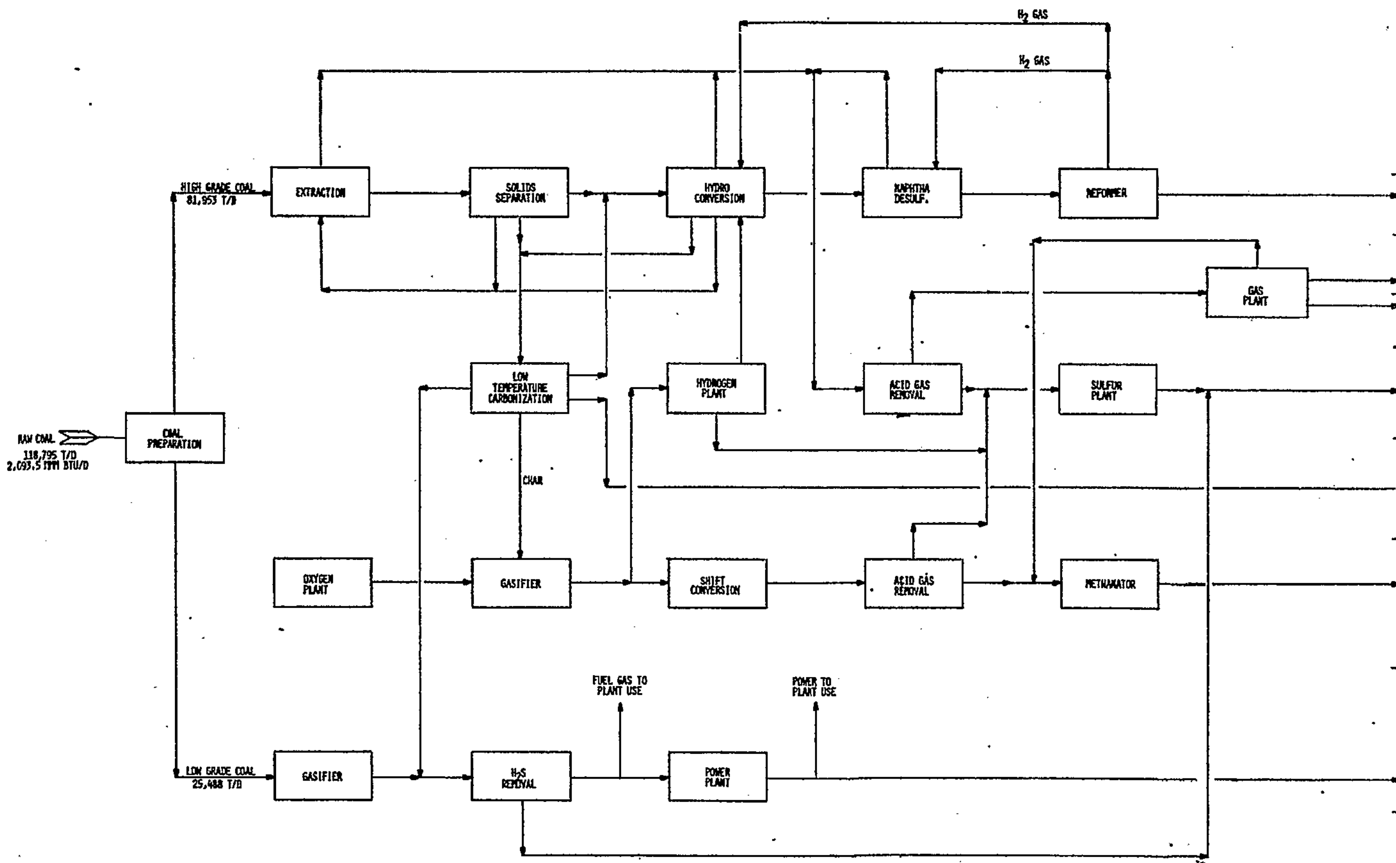
Table 4-1 - Pyrolysis Process Comparison

Item	Pressurized Flash Pyrolysis	Coalcon	High-Pressure Hydropyrolysis (1,000 psi)
	$ MM	$ MM	$ MM
Investment			
Pyrolysis unit (PU)	14.9	17.9	37.5
Land @ 1.76% PU	0.3	0.3	0.7
Working capital @ 3.83% PU	0.6	0.7	1.4
Total	15.8	18.9	39.6
	$ MM/Yr	$ MM/Yr	$ MM/Yr
Required income after tax (10% ROI)	1.6	1.9	4.0
Tax	1.7	2.1	4.3
Required income before tax	3.3	4.0	8.3
Expenses			
Utilities	14.3	18.5	46.0
Hydrogen @ $0.50/M SCF	--	25.5	40.3
Coal @ $14.00/ton	50.7	53.5	40.7
Total	65.0	97.4	127.0
Depreciation	1.2	1.4	2.9
Income			
Gross required income	69.5	102.8	138.2
Char credit @ $5/ton	6.2	6.2	5.2
Required gas + liquid credit	63.3	96.6	133.0
Gas + liquid cost $/MM Btu	1.49	1.70	1.95
Ratio	1.0 (base)	1.14	1.31

Table 4-2 - POGO Short List Candidate Comparison

Item	O/G Reference	F-T Reference	Case I	Case II	Case III	Case IV
Material Balance						
ROM coal, TPD (12% H_2O)	-	-	118,795	100,289	83,880	-
Feed coal, TPD (2.7% H_2O)	35,670	30,000	-	-	-	115,000
Coal HHV, Btu/lb	12,125	12,550	11,337	11,886	11,242	12,122
Process configuration	Oil/Gas	Fischer-Tropsch	CSF	Flash pyrolysis, CSF, and Fischer-Tropsch	Flash pyrolysis and Fischer-Tropsch	Flash pyrolysis and SRC
Type of gasifier	Entrained	Entrained	Entrained	Entrained	Entrained	Entrained
Products, percentage of feed coal energy						
Gas	20.1	35.6	12.3	17.2	19.9	15.6
LPG	4.7	1.9	4.9	5.0	4.8	3.0
Naphtha	6.3	13.0	9.2	7.2	6.6	14.1
Chemical naphtha	-	-	-	5.0	6.4	-
Jet fuel	-	-	-	2.2	2.8	-
Diesel	-	11.3	-	0.6	0.8	-
Fuel oil	45.2	3.8	35.5	22.5	17.6	28.6
Sulfur	1.1	1.5	1.2	1.1	1.2	1.2
Power	-	1.1	8.9	10.1	12.7	8.6
Alcohol	-	4.6	-	0.6	0.8	-
Ammonia	0.2	-	-	-	-	-
Total*	77.6	72.8	72.0	71.5	73.6	71.1
Costs Relative to Case IV						
Relative required product selling price of fuel product per Btu	1.01	1.32	1.05	1.07	1.10	1.00
Relative fixed capital investment per Btu product	1.27	1.98	1.28	1.30	1.33	1.00
Relative fixed capital investment per daily ton of coal fed	1.38	1.98	1.22	1.30	1.28	1.00

*Represents projected plant thermal efficiency.

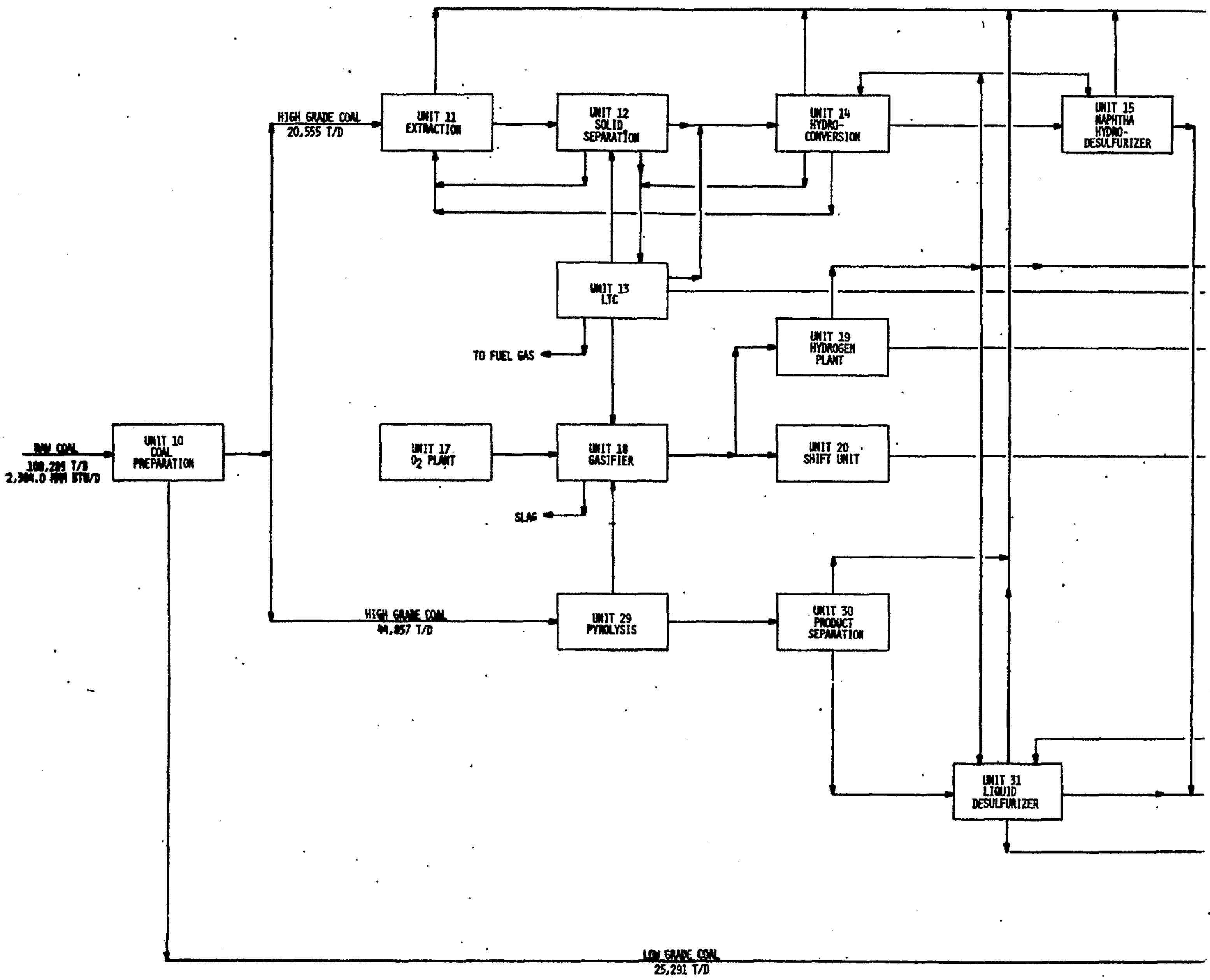


	T/D	MMSCFD OR BPD	MMMBTU/D
GASOLINE	6184.	46,623 BPD	249.2
C_3 LPG	1659.	18,456 BPD	71.9
C_4 LPG	1398.	13,667 BPD	59.6
SULFUR	4057.		32.3
HEAVY FUEL OIL	25,765.	147,742 BPD	956.2
SNG	7252.	336. MMSCFD	330.4
POWER 1000 MW			240.0
		TOTAL	1939.6
		EFFICIENCY	72.0%

Figure 4-1 POGO Plant Block Flow Diagram, Case I - CSF Process

A

B



HIGH GRADE COAL
20,555 T/D
UNIT 11
EXTRACTION
UNIT 12
SOLID
SEPARATION
UNIT 14
HYDRO-
CONVERSION
UNIT 15
NAPHTHA
HYDRO-
DESULFURIZER
UNIT 13
LTC
TO FUEL GAS
UNIT 19
HYDROGEN
PLANT
UNIT 10
COAL
PREPARATION
UNIT 17
O_2 PLANT
UNIT 18
GASIFIER
UNIT 20
SHIFT UNIT
SLAG
HIGH GRADE COAL
44,857 T/D
UNIT 29
PYROLYSIS
UNIT 30
PRODUCT
SEPARATION
UNIT 31
LIQUID
DESULFURIZER
LOW GRADE COAL
25,291 T/D

Figure 4-

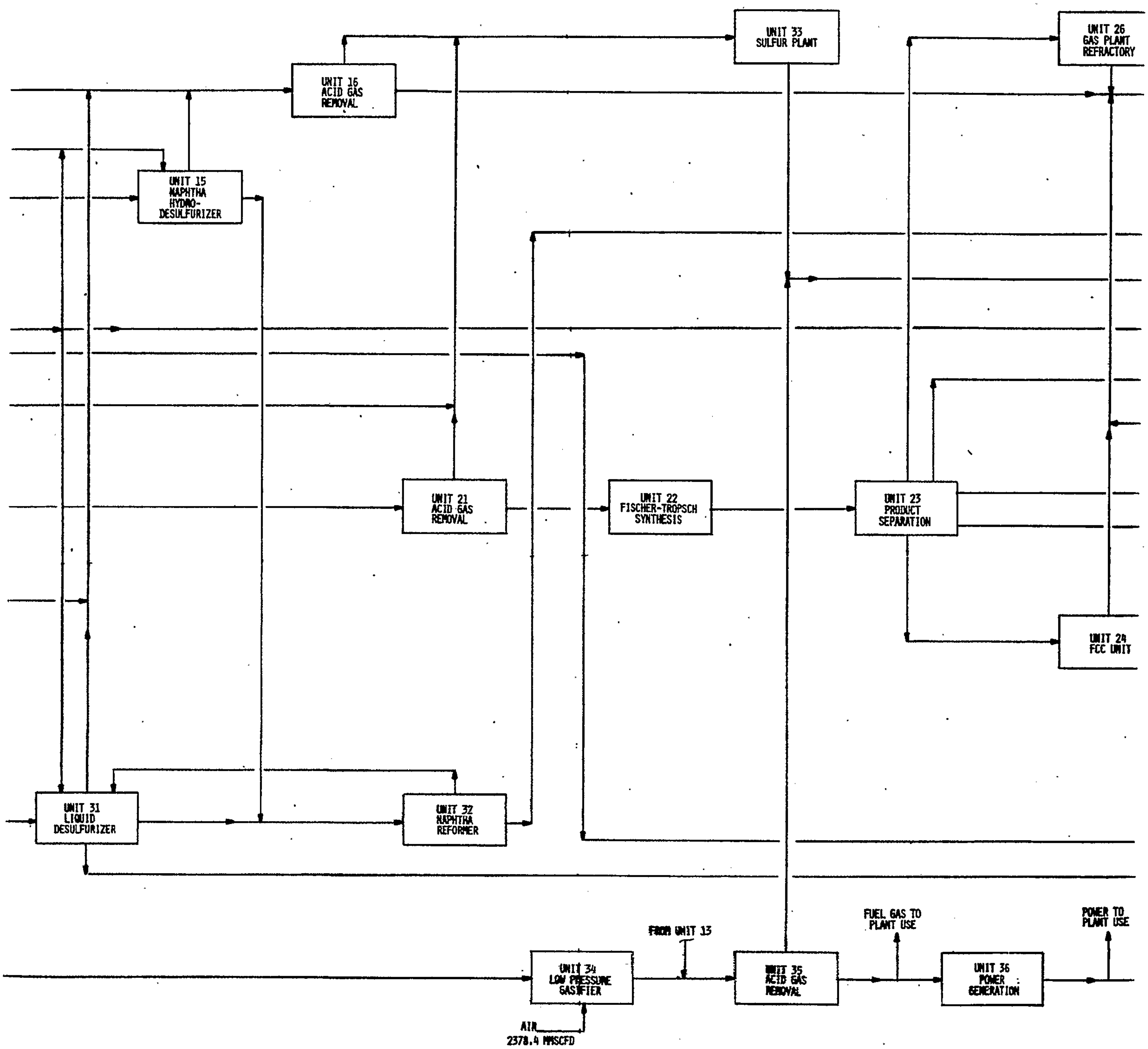
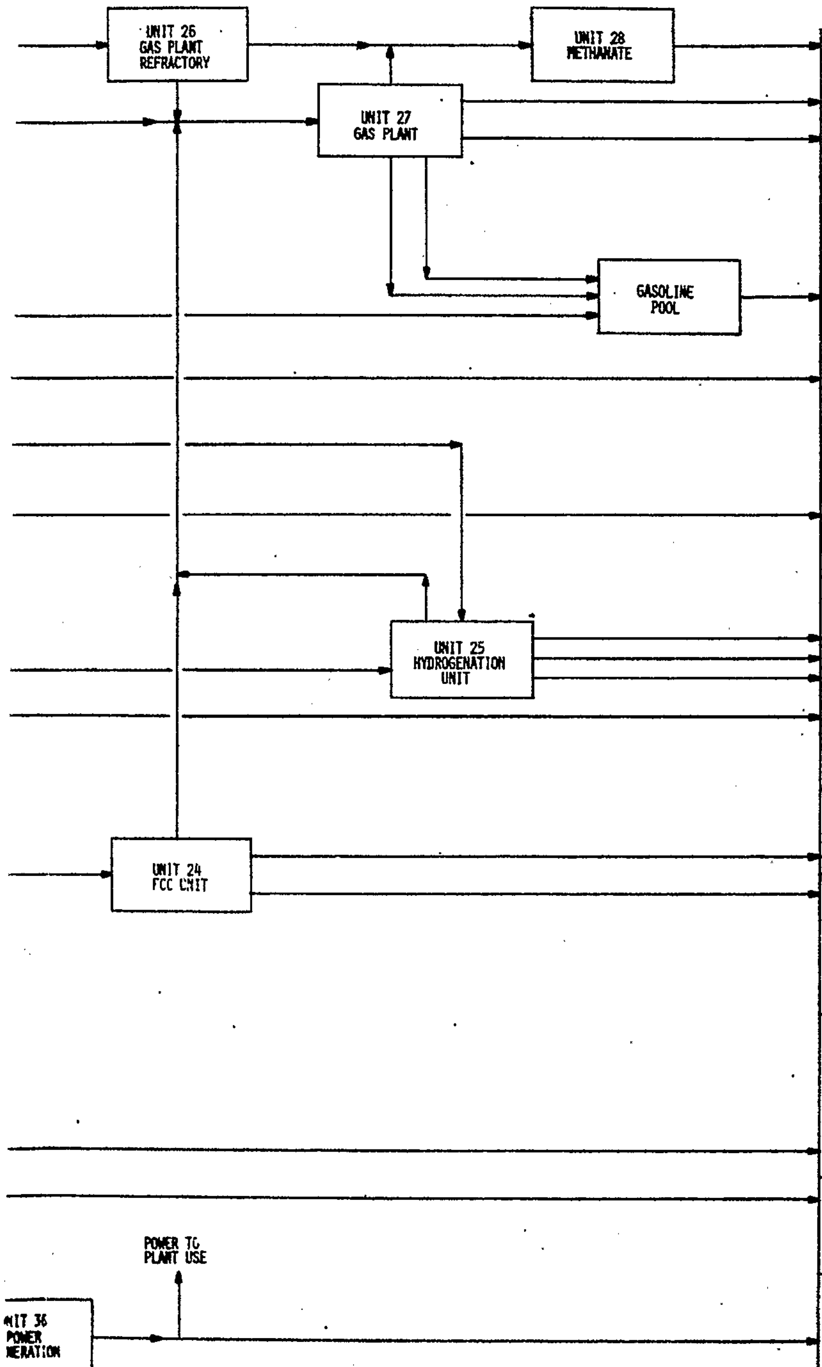
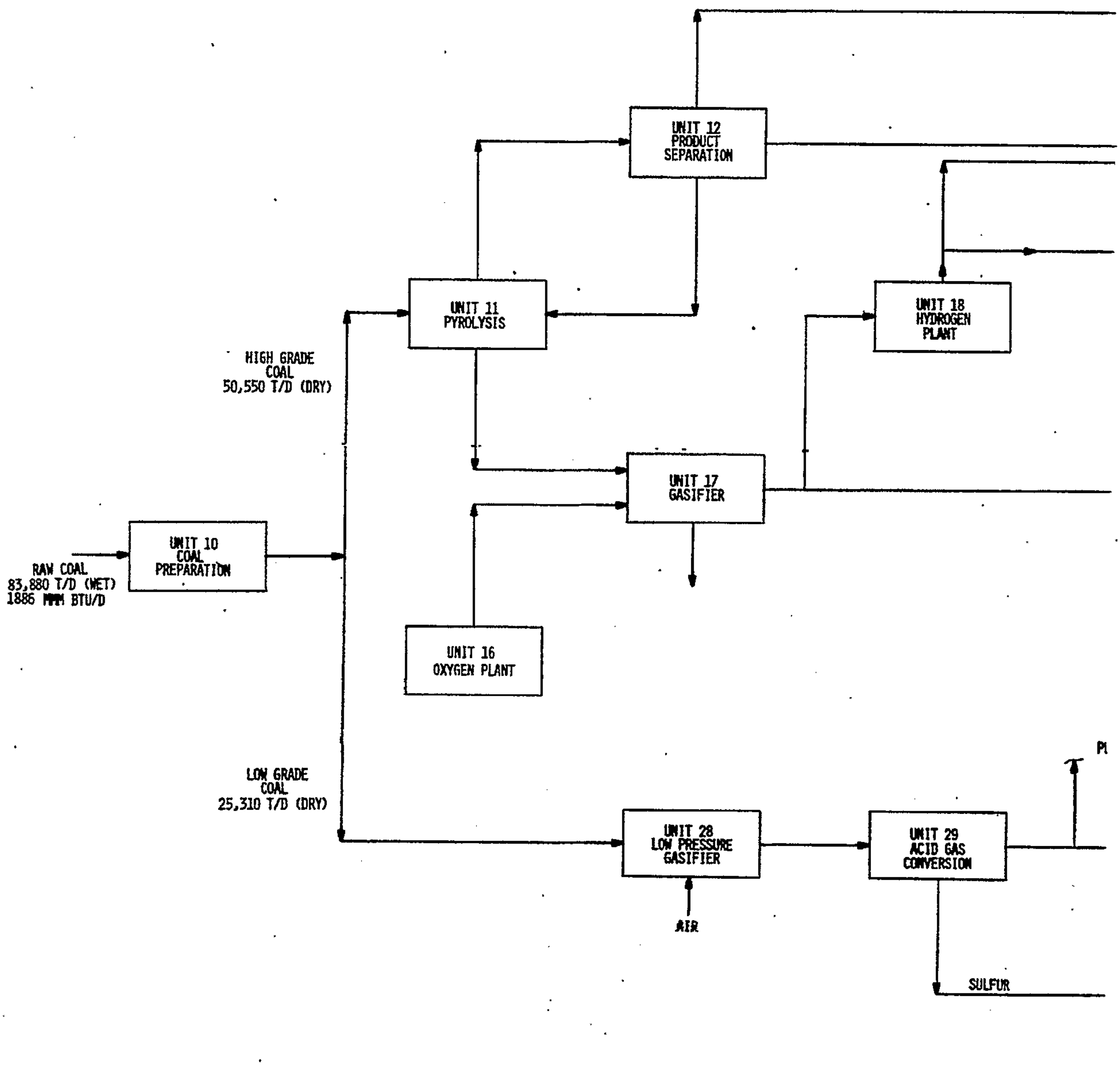


Figure 4-2 POGO Plant Block Flow Diagram, Case II - CSF Process, Flash Pyrolysis, and Fischer-Tropsch

B



	TONS/DAY	MMSCFD/BPD	MMM BTU/D
SNG	9,044	388 MMSCFD	408.4
PROPANES	1,238	13,773 BPD	53.6
BUTANES	1,556	15,211 BPD	66.3
POOL GASOLINE	4,382	29,706 BPD	172.3
SULFUR	3388	—	27.0
OXYGENATES	566	4,097 BPD	15.0
LT. CHEMICAL NAPHTHA	1,202	10,512 BPD	120.0
HEAVY CHEMICAL NAPHTHA	1,724	13,978 BPD	
JET FUEL	1,294	9,862 BPD	52.0
DIESEL	264	1,908 BPD	15.0
DIESEL	122	802 BPD	
HEAVY FUEL OIL	60	305 BPD	2.3
HEAVY FUEL OIL	6,462	37,056 BPD	239.8
HEAVY FUEL OIL	7,696	46,723 BPD	292.6
POWER 1,000 MW			240.0
		TOTAL	1704.3
		EFFICIENCY	71.5%



UNIT 12
PRODUCT
SEPARATION
UNIT 11
PYROLYSIS
UNIT 18
HYDROGEN
PLANT
HIGH GRADE
COAL
50,550 T/D (DRY)
UNIT 17
GASIFIER
UNIT 10
COAL
PREPARATION
RAW COAL
83,880 T/D (WET)
1886 MMM BTU/D
UNIT 16
OXYGEN PLANT
LOW GRADE
COAL
25,310 T/D (DRY)
UNIT 28
LOW PRESSURE
GASIFIER
AIR
UNIT 29
ACID GAS
CONVERSION
SULFUR

A

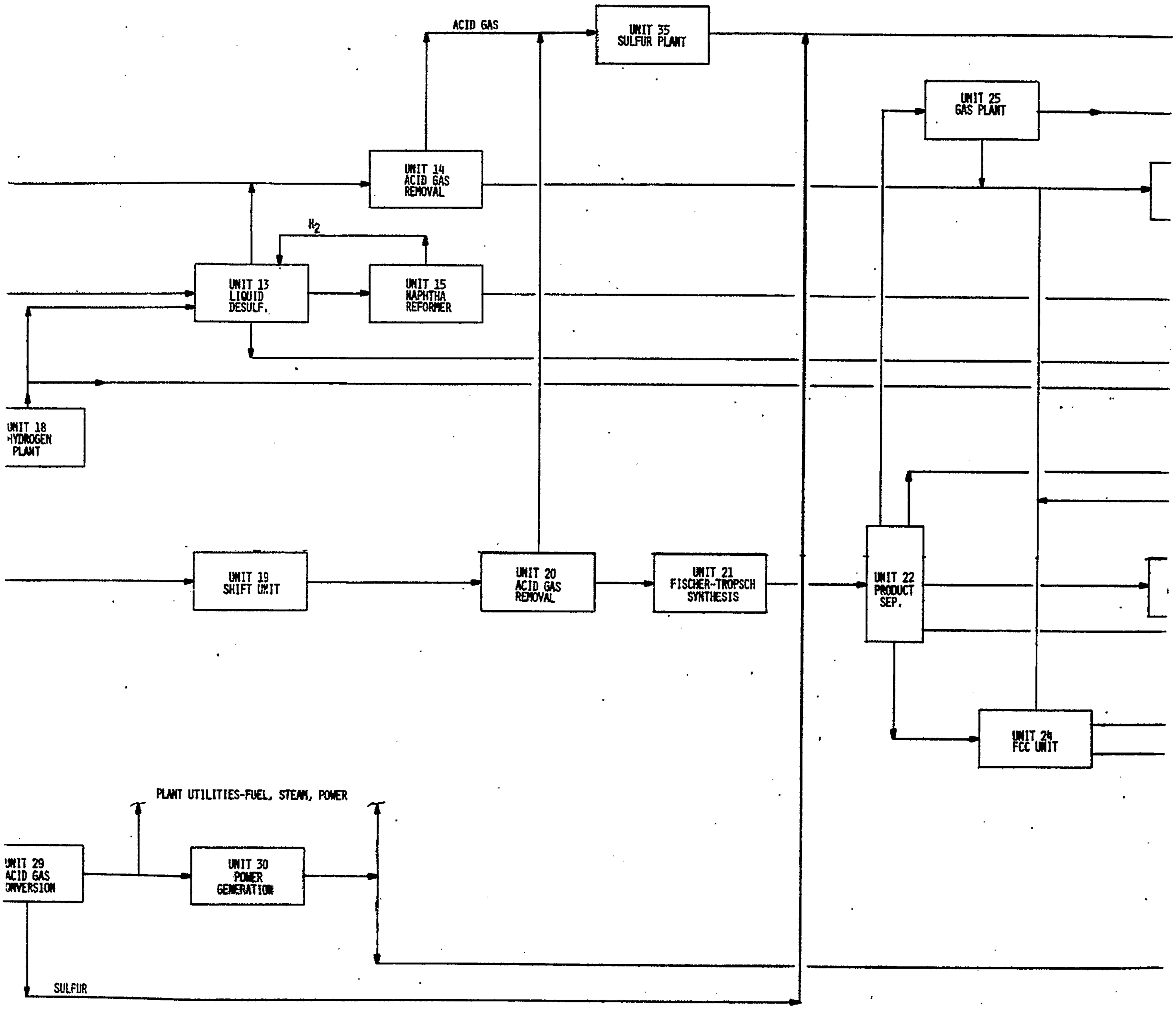


Figure 4-3 . POGO Plant Block Flow Diagram, Case III – Pyrolysis and Fischer-Tropsch

β

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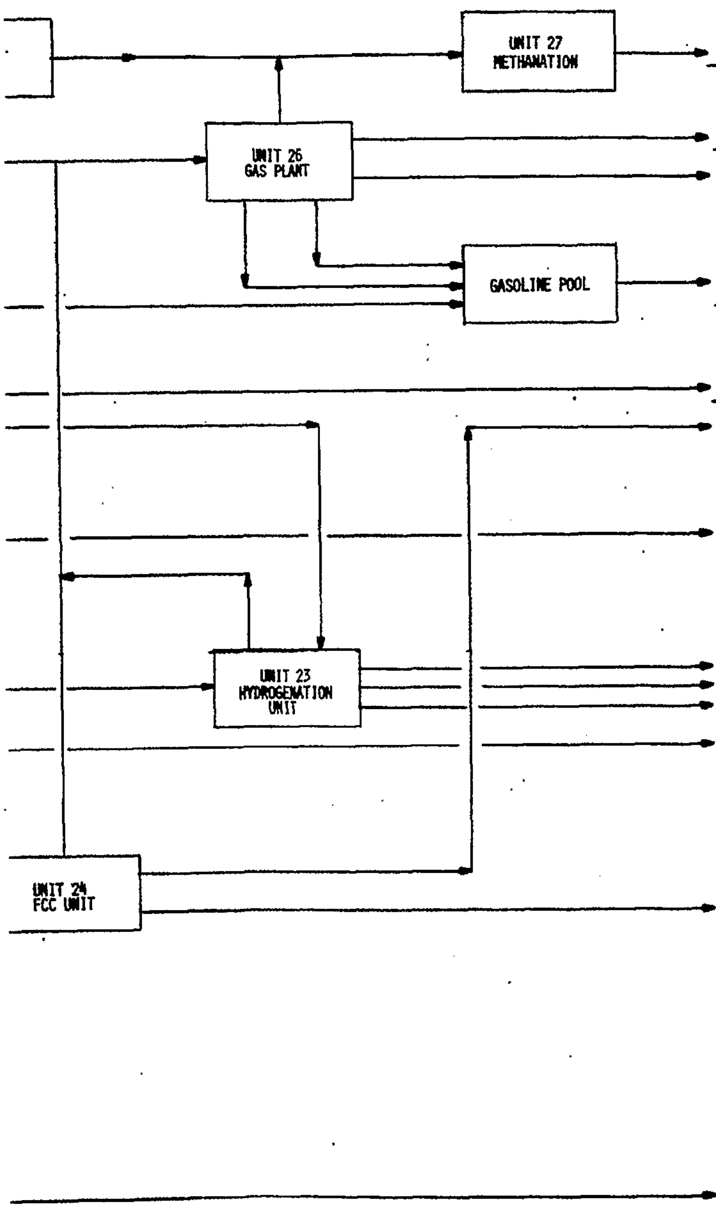

	TONS/DAY	BPD OR MMSCFD	MMM BTU/D
SULFUR	2,740		22
SNG	8,702	372 MMSCFD	376
PROPANES	847	9,980 BPD	39
BUTANES	1,207	11,802 BPD	51
POOL GASOLINE	3,145	21,428 BPD	124
HEAVY FUEL OIL	8,663	52,597 BPD	331
HEAVY FUEL OIL	49	249 BPD	
OXYGENATES	566	4,097 BPD	15
LIGHT CHEMICAL NAPHTHA	1,202	10,512 BPD	120
HEAVY CHEMICAL NAPHTHA	1,724	13,978 BPD	
JET FUEL	1,294	9,862 BPD	52
DIESEL	264	1908	15
DIESEL	122	802 BPD	
POWER 1000 MW			240

TOTAL 1385 (73.4% EFF.

C

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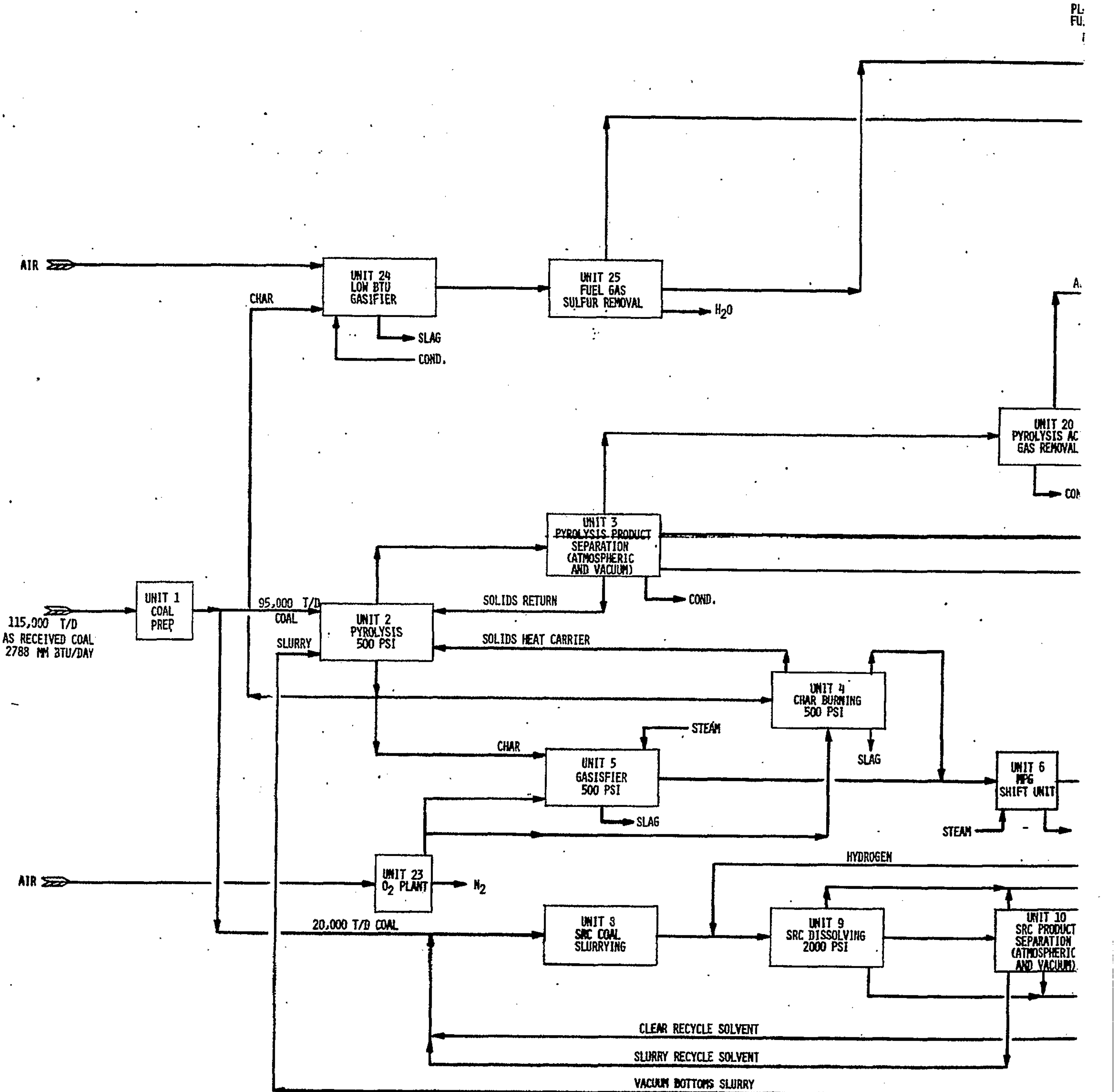

AIR
UNIT 24
LOW BTU
GASIFIER
CHAR
SLAG
COND.
UNIT 25
FUEL GAS
SULFUR REMOVAL
H2O
UNIT 20
PYROLYSIS AC
GAS REMOVAL
UNIT 3
PYROLYSIS PRODUCT
SEPARATION
(ATMOSPHERIC
AND VACUUM)
COND.
115,000 T/D
AS RECEIVED COAL
2788 MM BTU/DAY
UNIT 1
COAL
PREP
95,000 T/D
COAL
SLURRY
UNIT 2
PYROLYSIS
500 PSI
SOLIDS RETURN
SOLIDS HEAT CARRIER
UNIT 4
CHAR BURNING
500 PSI
SLAG
STEAM
CHAR
UNIT 5
GASISFIER
500 PSI
SLAG
UNIT 6
MPG
SHIFT UNIT
STEAM
AIR
UNIT 23
O2 PLANT
N2
HYDROGEN
20,000 T/D COAL
UNIT 8
SRC COAL
SLURRYING
UNIT 9
SRC DISSOLVING
2000 PSI
UNIT 10
SRC PRODUCT
SEPARATION
(ATMOSPHERIC
AND VACUUM)
CLEAR RECYCLE SOLVENT
SLURRY RECYCLE SOLVENT
VACUUM BOTTOMS SLURRY

A

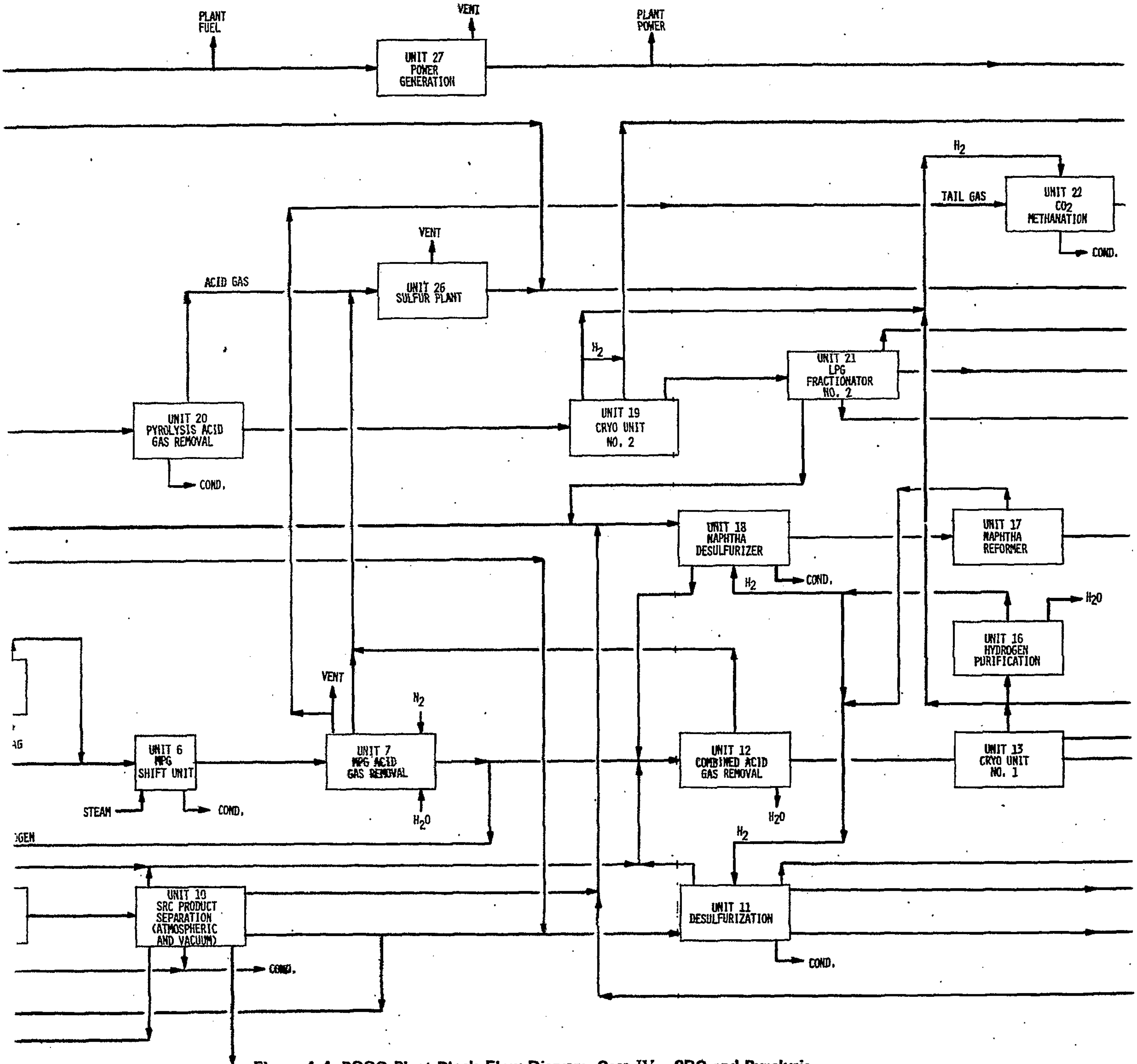
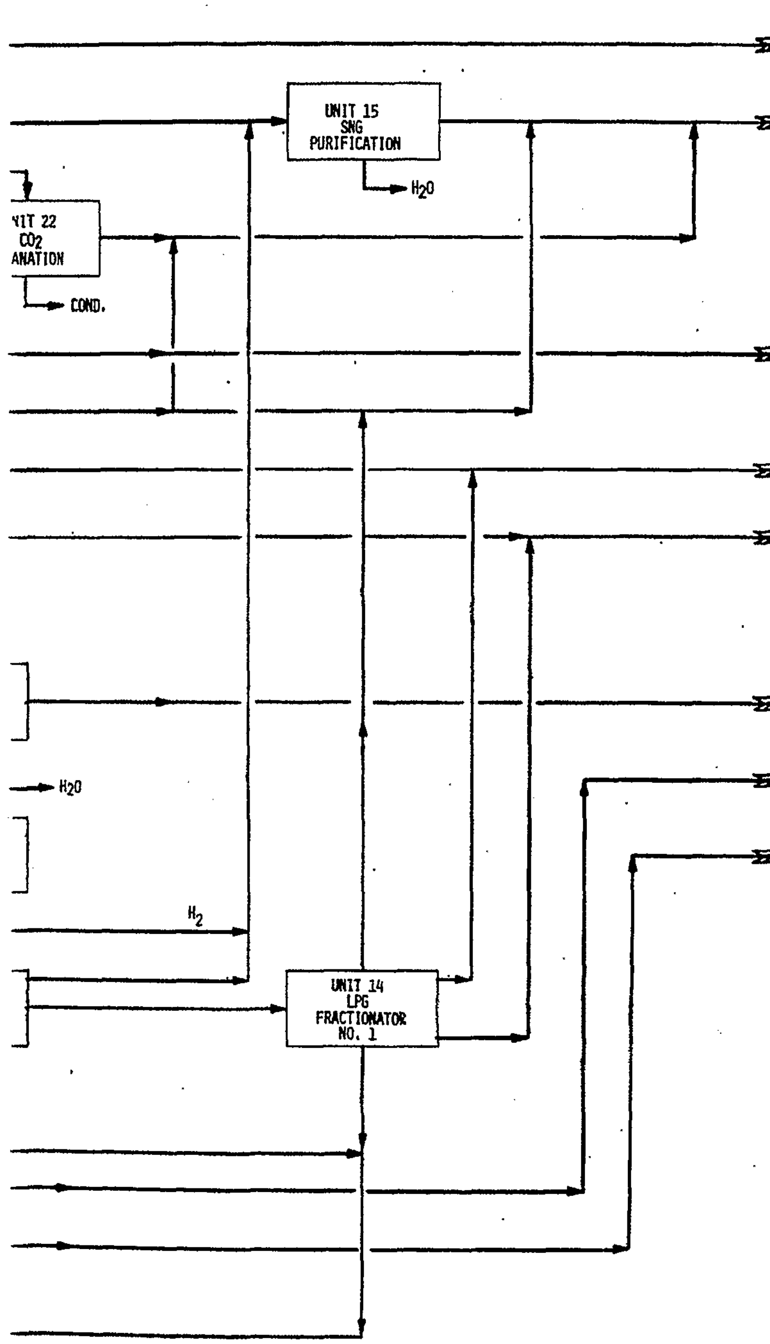


Figure 4-4 POGO Plant Block Flow Diagram, Case IV - SRC and Pyrolysis



	T/D	B/T OR MMSCFD	MMM BTU/D
POWER 1000 MW			240.
SNG	10,520	414	435
SULFUR	4157	———	33
C_3 LPG	975	11,005	43.
C_4 LPG	973	10,958	41.
GASOLINE	9824.	72,074	393.
DISTILLATE FUEL	15,400	103,610.	603.
HEAVY FUEL OIL	5476	28,769.	196.
TOTAL			1984.

THERMAL EFFICIENCY 71.2%

c

SECTION 5

PREFERRED PROCESS SELECTION

The information developed in the preceeding short list report section was used as a basis for analysis and selection of a preferred process configuration. The selection was guided by the projected technical and economic factors that were available at that time. In addition to the quantitative projections, the analysis included some subjective judgments regarding the current characteristics and projected potentials of the technologies, both when operated alone and in combination with other coal conversion processes. The results of these analyses and final configuration recommendations are summarized below.

RECOMMENDED CONFIGURATION

Results of the analysis indicate that Cases III and IV were defined to be the two prime candidates for the POGO design. The reasons included the fact that economic factors have indicated the SRC-type hydroliquefaction to be a potentially slightly lower cost producer than the CSF; it is also indicated to have a potential for future cost reduction when operated in combination with a companion coal conversion process such as flash pyrolysis. Also, the SRC-type liquefaction has been successfully operated for extended periods on large, pilot plant scale, while the CSF process must still demonstrate a similar successful experience.

Results of the preliminary screening effort indicate that the CSF process, by the nature of its characteristics, did not offer an equivalent potential for future cost reductions when operated in combination with a pyrolysis unit. The Fischer-Tropsch combination, Case III, while indicating a high projected RPSP on a dollars per Btu basis, appears to offer potential for increase in rates of return on capital because of the nature of its product characteristics. Its product, having nil sulfur, nitrogen, and particulate content, could potentially bring premium product prices for environmental reasons; also, the liquid hydrocarbons could serve as valuable petrochemical feedstocks.

Based on the results of the analysis, Case IV has been selected and recommended, with some modification, as the preferred design configuration. Advantages of this case include:

- Economic - A preliminary cost comparison indicated that this configuration may require approximately 40% lower capital investment per daily ton of coal and about 15% lower average product selling price at 12% DCF than Case III; see Table 4-2. The pyrolysis-SRC scheme also has the potential for further economic improvement based on a subsequent analysis effort - these improvements include the use of beneficiated coal feed and improved thermal efficiency.

- Thermal Efficiency - Studies of the SRC process and the Fischer-Tropsch process have shown that the SRC process has the potential to operate at an overall process thermal efficiency several percentage points greater than the Fischer-Tropsch process.

- Elimination of Filtration - A difficult and expensive process step in the SRC process is the separation of unreacted coal and ash from liquified coal products. By combining pyrolysis of coal with pyrolysis of SRC-vacuum distillation residue in the SRC II mode of operation, the use of filters or other solid separation devices is eliminated. The majority of the heavy residue is recovered as salable light liquids or gases.

- Plant Integration - The design has potential to achieve high thermal efficiency by integration of the plant utilities and process units. The SRC process and supporting units in the recommended configuration are expected to have a slightly greater flexibility than the flash pyrolysis/Fischer-Tropsch combination.

- Technical Background - The basic hydroliquefaction technology that is the genesis of the SRC process is a proven industrial operation, both in German plants operated during World War II and in two pilot plants now in operation in the United States.

CONCLUSION

Comparison of candidate processing schemes has led to the selection of the Case IV flash pyrolysis/SRC-based scheme as the recommended configuration. It is judged that the configuration process can be further improved.

The proposed preliminary design criteria for the conceptual POGO design is presented in Section 6. This criteria should define the expected characteristics for the user and permit the designers to proceed with their work.

SECTION 6

PRELIMINARY DESIGN CRITERIA

The preceding process selection led to the preparation of a preliminary design criteria to describe the proposed POGO conceptual commercial plant at the time of start of the directed process design. This section presents the document: "POGO Conceptual Commercial Plant, Preliminary Design Criteria," ERDA Contract No. (49-18)-1775. Improvements and revisions developed during the course of the directed process design are to be incorporated into the final design basis document.

DESIGN OBJECTIVES

The objective is a commercial POGO design; this will be a conceptual coal-oil-gas (COG) refinery to produce power, oil, gas, and other products (POGO). The results of preliminary studies indicate that the complex should include SRC II and pyrolysis process units, product-finishing units, and utilities production. The utilities section will include production of 1,000 MW of export power with the objective of a minimum fixed capital investment (FCI) requirement and production costs.

The preliminary design is to be in sufficient detail for a fixed capital estimate with target accuracy of -5 + 20% estimate and profitability analysis.

DESIGN PARAMETERS

Five factors are considered in the design parameters.

(1) Design Capacity: 20,000 TPD of coal charged to dissolver.

Pyrolysis unit shall be charged coal of sufficient capacity to provide the char required to generate the required syngas.

(2) Site Location: Preliminary designs will be developed for three separate locations:

- Eastern region of the U.S. interior coal province
- Lower Appalachia
- Rocky Mountain coal province

(3) Coal Feed: Illinois No. 6 seam coal.

Requirements of cleaned coal are to supply feed to dissolver unit, pyrolysis unit, and fuel gas. The pyrolysis unit produces char, which will serve as a gasifier feed for conversion to syngas and hydrogen gas.

The coal properties include cleaned, sized, and dried Illinois No. 6 seam coal with the following typical analysis:

(a) Proximate analysis (wt % of cleaned, dried coal)

Moisture	2.7
Ash	11.8
Volatile matter	39.7
Fixed carbon	45.8

Gross heating value 12,125 Btu/lb

(b) Ultimate analysis (wt % MAF basis)

Carbon	78.6
Hydrogen	5.4
Nitrogen	1.5
Sulfur	4.3
Oxygen	10.2

(4) Coal will be produced in a captive mine.

(5) Products: Liquid products will consist of LPG, gasoline, distillate, and heavy fuel oils.

SNG of pipeline quality with a heating value in the range of 950 to 1,050 Btu/SCF.

Byproduct sulfur, ammonia, and electricity will be produced. Product slate expected is:

LPG

Gasoline pool

Diesel fuel

Fuel oil

SNG

Ammonia (if justified)

Sulfur

Power

POGO PROCESS PLANT DESCRIPTION

This section describes current thinking regarding major equipment and process elements of the plant. Changes will be made during the design development as appropriate to achieve the stated objectives.

A block flow diagram, Figure 4-4, depicts the anticipated processing sequence. All effluents are to meet environmental standards.

(A) Dryer

Dryers will be used to remove free water and a small portion of the inherent moisture.

(B) Grinders

Dried 1-1/4 in. x 0 coal will be ground to produce a pyrolyzer feed.

(C) Flash Pyrolysis Unit

Coal will be bed to the system by a dry feeder system. The pyrolysis unit will include its own heat input system. Solids slurry from product separation will be fed to the pyrolysis zone.

Products from the pyrolysis unit will consist of:

- Char as feed to gasifier
- Gas
- Distillate

(D) Pyrolysis Products Separation

This will provide a quench (and heat recovery) section and facilities to separate gas and liquids. Gas will be directed to the purification unit and the gas plant. Liquid treatment will include particulate matter removal and fractionation as required to produce desulfurized unit feed.

Facilities for any required gas recycle for pyrolysis operation are included in this unit.

(E) Gasifier

This design will use the following features:

(1) A steam-oxygen entrainment-type gasifier will be used to produce syngas.

(2) An air or oxygen entrainment-type gasifier will be used to produce fuel gas; final selection will be based on the results of additional studies.

(F) Gas-Solids Removal

The char solids entrained in the gasifier product gas stream are to be removed by such means as high efficiency cyclones, venturi scrubbers, electrostatic precipitators, and wash columns.

(G) Acid Gas Cleanup

CO_2 and H_2S are to be removed by scrubbing with a physical solvent in a gas treating unit. A clean CO_2 stream is discharged to the atmosphere. An H_2S/CO_2 mixture is directed to the sulfur recovery unit.

(H) Shift Conversion

This unit will convert most of the CO to H_2. A selective acid gas removal unit will be used to remove CO_2 and H_2S in two streams: One, CO_2 in sufficient purity for discharge to the atmosphere, and the other, a mixture of H_2S in CO_2 sufficiently concentrated in sulfur to make a good sulfur plant feed.

(I) Dissolving Unit

Coal will be slurried with a solvent consisting of 2/3 slurry and 1/3 filtrate solvent. Total solvent-to-coal weight ratio will be 3:1. Coal slurry will be contacted with hydrogen gas at about 2,000 psi total pressure in the dissolver, with a slurry residence time of 15 minutes.

(J) Dissolver Products Recovery

Product slurry will be separated from recycle gases and then flashed in several stages. Wherever economically attractive, power will be recovered from gas expanders and pressure letdown turbines. An atmospheric tower will separate noncondensibles, naphtha, wash oil and distillate oil, and slurry recycle solvent and vacuum tower feed. The vacuum tower will concentrate net-heavy liquid product slurry so that vacuum bottoms containing solids can be fed to the pyrolyzer for liquid recovery or conversion to char.

(K) Gas Treating

Off-gas from the dissolver will be treated in an amine-type acid gas removal unit to remove CO_2 and H_2S, and then it will be sent to a cryogenic separation unit to remove methane and heavier hydrocarbons from H_2 and CO. Purified H_2 and CO will then be recycled to the dissolver preheater. The methane will be separated from the LPGs, and purification units will produce the specification SNG, propane-LPG, and butane-LPG.

(L) Sulfur Recovery

Sulfur will be recovered from the process and utility treating units H_2S gas effluent.

(M) Naphtha Reformer

A naphtha reformer shall be provided as required to produce a pool gasoline whose research octane number is 96.0.

(N) Naphtha Desulfurizer

Recovered naphthas shall be catalytically hydrogenated to convert sulfur values to H_2S and nitrogen values to NH_3 for removal. Severity of operation shall be that required to produce acceptable naphtha-reformer feedstock.

(O) Distillate Desulfurization

Catalytic desulfurization of distillate products shall be accomplished to produce acceptable distillate fuels or diesel products.

(P) Fuel Gas Production

Char or coal will be fed to an entrained gasifier to produce sufficient fuel gas for all plant fuel needs such as steam and power generation, plus heater firing. Additional studies will determine the final gasifier design and configuration. Alternatives to be studied include type of oxidant (air or oxygen), type of carbonaceous feed (char or coal), pressure, temperature profile, slag removal, and mechanical design configurations.

(Q) Water and Waste Gas Treating

All contaminated plant water streams will be collected and treated to remove dissolved gases. The gases, consisting mainly of NH_3 and H_2S, will be separated to produce anhydrous NH_3, a salable product, and H_2S for feed to the sulfur plant. The sulfur plant will, in turn, convert the H_2S to elemental sulfur and a clean stack gas. Stripped water will be sent to process use, where dissolved hydrocarbons will be destroyed.

(R) Oxygen Plant

Oxygen shall be produced in commercial-type oxygen plants using whatever economic head and/or material sources are available in the operating plant or utility section.

(S) Power Generation Unit

A utility-type unit shall produce steam and electric power for use in the processing plant, in addition to 1,000 MW of electric power for sale. This plant will be capable of continuous output in the same manner as public utilities must.

SECTION 7

LITERATURE CITED
POTENTIALS FOR EMERGING ENERGY TECHNOLOGIES

1. O'Hara, J.B., Jentz, N.E., Rippee, S.N., and Mills, E.A., "Demonstration Plant - Clean Boiler Fuels from Coal - Preliminary Design," Volume I - Summary Report, Volume II - Capital Cost Estimate, Volume III - Economic Analysis, OCR R&D Report No. 82, Interim Report No. 1, 1973.

2. O'Hara, J.B., Crawford, C.L., Hincks, H.F., and Mills, E.A., "Commercial Complex, Conceptual Design/Economic Analysis, Oil and Power by COED-Based Coal Conversion," ERDA R&D Report No. 114, Interim Report No. 1, 1974.

3. O'Hara, J.B., Hervey, G.H., Fass, S.M., and Mills, E.A., "Oil/Gas Plant Design Criteria," CEP Capsule, Chemical Engineering Progress, Vol. 72, No. 8, pp. 78-79, August 1976.

4. O'Hara, J.B., Bela, A., Jentz, N.E., and Khaderi, S.K., "Fischer-Tropsch Plant Design Criteria," CEP Capsule, Chemical Engineering Progress, Vol. 72, No. 8, pp. 65-67, August 1976.

5. Coles, E.T., and Crebar, C.R., "Coalcon, Commercial Plant Process Evaluation Report," ERDA Report, Contract E(49-18)-1736.

6. Dravo Corporation, "Conceptual Commercial Design and Commercial Feasibility Evaluation for Clean Boiler Fuel Facility," ERDA Report, Volume I - Conceptual Commercial Design, Volume II - Sections and Evaluations, Volume III - Commercial Feasibility.

7. O'Hara, J.B., et al., "Fischer-Tropsch Complex, Conceptual Design/Economic Analysis," ERDA R&D Report No. 114, Interim Report No. 3, January 1977.

8. O'Hara, J.B., et al., "Oil/Gas Complex, Conceptual Design/Economic Analysis," ERDA R&D Report No. 114, Interim Report No. 4, March 1977.

9. Lowry, H.H., "Chemistry of Coal Utilization - Supplementary Volume," John Wiley and Sons, N.Y., p. 356, 1963.

APPENDIX A

PROCESS DESCRIPTIONS AND BALANCES

SRC I AND SRC II PROCESS ALTERNATIVES

Data Sources and Status

The data was supplied by the Pittsburgh and Midway Coal Mining Company (P&M) for the coal liquefaction process and yields. This consisted of progress reports prepared under their ERDA Contract E(49-18)-496, plus certain information exchanges during the course of Parsons work for ERDA as a Technical Evaluation Contractor under Contract E(49-18)-1234.

Basic data and yields for the gasification units originated from work done for the Office of Coal Research (OCR) by Bituminous Coal Research, Inc. (BCR) under Contract 14-32-0001-1207. Independent designs were developed by Parsons for the specific conditions of temperature, pressure, and gas composition selected for this process assessment work.

Development of the SRC process began in 1962 at the Spencer Chemical Company. The work was later transferred to P&M, which is a subsidiary of Gulf Oil Corporation. At present, two pilot plants are operating: one with a 50-TPD coal feed capacity at Tacoma, Washington, by P&M, and the other with a 6-TPD capacity at Wilsonville, Alabama, by Southern Services Incorporated.

The Wilsonville plant began operation in February, 1974, and has produced specification grade solvent refined coal (SRC) using several coals. The Tacoma plant was completed in the fall of 1974 and has been operated for over 1,500 hours, since early 1975. Most of the operation has been in the SRC I mode. However, five heat and material balance runs were made in the SRC II mode. Extensive work has been done at the P&M Merriam, Kansas, process development unit in the SRC II mode.

Process Description

The SRC process includes a coal liquefaction section developed by P&M and a 2-stage gasification unit with characteristics similar to the unit under development by BCR and the ERDA Bi-Gas pilot plant located at Homer City, Pennsylvania.

The SRC process is depicted in the Figure A-1 block flow diagram.

A 1/8-inch, minus, feed coal is combined with solvent to form coal slurry, which is pumped to the preheat furnace. Syngas or hydrogen is added at the furnace entrance, the resulting mixture is preheated, and then it is fed to the dissolver, which is operated at about 850°F and 2,000 psig.

The product mixture from this reaction system consists of a liquid phase, a solid phase of ash plus undissolved coal, and a gas phase. The gas phase is separated, scrubbed to remove H_2S and CO_2, and its major portion is combined with makeup syngas or hydrogen and recycled to the preheat furnace inlet. The excess gas is further processed for sulfur removal prior to ejection to the atmosphere. The solid phase is separated from the liquid phase by filtration, it is dried with solvent recovery, and then transferred to the gasification plant, where the residual carbonaceous material is gasified to produce syngas.

The SRC I mode of operation uses a distillate process solvent that is produced by vacuum distillation and followed by fractionation. A solid deashed product with a sulfur content in the range of 0.8 to 0.9% is produced. In the SRC II mode of operation, a portion of the unfiltered dissolver liquid product, containing undissolved coal particles and ash, is used for recycle to slurry the feed coal. This results in a higher ash content in the dissolver providing a pseudocatalytic effect, a longer average retention time, and a high, hydrogen-to-carbon ratio liquid product with a lower sulfur content (about 0.4 to 0.6%).

The yield data for each of the four SRC modes was simulated by a proprietary computer program. Material balances for the separate modes are indicated on block flow diagram for each mode:

Mode	Figure No.
SRC I - H_2 gas	A-1
SRC I - Syngas	A-2
SRC II - H_2 gas	A-3
SRC II - Syngas	A-4

In each mode, equipment has been provided to produce 1,000 Btu/SCF pipeline gas. This plant is self-contained, producing all of its own power and fuel gas. The only utility required is fresh makeup water. A low Btu gasifier has been added to the original clean boiler fuel design (Ref. 1) to produce the fuel gas required for the plant operation. The dry filter cake produced in the SRC unit is combined with additional fresh coal and consumed in the low Btu gasifier.

Heat and Material Balance Factors

Key factors for the four SRC operational modes are:

SRC I - H_2 gas

Raw water usage	10,000 gal/min
ROM coal charged	51,500 TPSD
Washed & dried coal charged	36,200 TPSD
Heating value of fuel products	680 MMM Btu/SD HHV

SRC I - Syngas

Raw water usage	11,500 gal/min
ROM coal charged	52,300 TPSD
Washed & dried coal charged	36,900 TPSD
Heating value of fuel products	680 MMM Btu/SD HHV

SRC II - H_2 gas

Raw water usage	12,200 gal/min
ROM coal charged	55,900 TPSD
Washed & dried coal charged	39,340 TPSD
Heating value of fuel products	717 MMM Btu/SD HHV

SRC II - Syngas

Raw water usage	13,800 gal/min
ROM coal charged	57,580 TPSD
Washed & dried coal charged	40,550 TPSD
Heating value of fuel products	735 MMM Btu/SD HHV

Additional heat and material balance factors for hydrogen consumption include:

	H_2 Consumption Wt % of MAF Coal
SRC I - H_2 gas	2.24
SRC I - Syngas	2.24
SRC II - H_2 gas	3.33
SRC II - Syngas	3.33

Feeds

	TPSD	MMM Btu/SD HHV
SRC I - H_2 gas		
Coal to dissolver	25,000	630
Coal to gasifier	6,300	160
Coal to low Btu gasifier	4,900	120
Total	36,200	910
SRC I - Syngas		
Coal to dissolver	25,000	630
Coal to gasifier	6,600	170
Coal to low Btu gasifier	5,300	130
Total	36,900	930
SRC II - H_2 gas		
Coal to dissolver	25,000	630
Coal to gasifier	8,730	220
Coal to low Btu gasifier	5,610	140
Total	-39,340	990

	TPSD	MMM Btu/SD HHV
SRC II - Syngas		
Coal to dissolver	25,000	630
Coal to gasifier	9,550	240
Coal to low Btu gasifier	6,000	150
Total	40,550	1,020

Products

SRC I - H_2 gas	TPD	MMM Btu/SD HHV
Pipeline gas	2,600	120
LPG	390	20
Naphtha	860	40
SRC (400°+) 0.47 wt % S	16,000	500
Sulfur	1,130	–
Total	20,980	680

Indicated thermal efficiency, 75%

SRC I - Syngas		
Pipeline gas	2,800	130
LPG	240	10
Naphtha	860	40
SRC (400°+) 0.47 wt % S	16,000	500
Sulfur	1,150	–
Total	21,050	680

Indicated thermal efficiency, 73%

	TPD	MMM Btu/SD HHV
SRC II - H_2 gas		
Pipeline gas	3,910	180
LPG	240	10
Naphtha	700	30
SRC (400°+) 0.47 wt %	14,900	500
Sulfur	1,240	—
Total	20,990	720

Indicated thermal efficiency, 73%

	TPD	MMM Btu/SD HHV
SRC II - Syngas		
Pipeline gas	4,380	200
LPG	70	5
Naphtha	700	30
SRC (400°+) 0.47 wt %	14,900	500
Sulfur	1,280	—
Total	21,330	735

Indicated thermal efficiency, 72%

Estimated Operating Labor

	Operators/Shift
SRC I - H_2 gas	65
SRC I - Syngas	65
SRC II - H_2 gas	63
SRC II - Syngas	63

H-COAL

Data Sources and Status

This study was based on data developed and reported in progress reports by Hydrocarbon Research Institute, Inc. (HRI), under OCR Contract 14-01-0001-477. The FCI estimate for the dissolving, hydrogenation, and distillation sections were obtained from the report, "Evaluation of Project H-Coal," by American Oil Company published under OCR Contract No. 14-01-0001-1188, December 8, 1967, and escalated to 1975.

This report was based on the use of Illinois No. 6 coal of slightly different composition from that used in the present work. Yields were adjusted to make the H-Coal process consistent with those in this report.

An extension of the H-Coal process study will require additional pilot plant data.

This process is a related application of the ebullated-bed, H-Oil process previously developed by HRI and Cities Service Oil Company to convert heavy oil residues into lighter fractions. The OCR sponsorship of the H-Coal process began in 1965. A bench scale unit to process 25 pounds of coal per day has been operated for approximately 1,200 on-stream days. This unit has a reactor with a diameter of 0.8 inch and a height of 7 feet.

A process development unit with a capacity to process 3 TPD of coal has been operated with continuous runs of a 60-day duration. The process development unit has a reactor with a diameter of 8 inches and a height of 22 feet.

Design work is presently under way for a 600-TPD pilot plant.

Process Description

A block flow diagram presenting the major process steps and material balance is shown in Figure A-5. The hydroliquefaction section is the heart of the plant. Here, a slurry of pulverized coal in process-produced heavy gas-oil is pumped by special slurry pumps to about 3,000 psig, where it is mixed with hydrogen, preheated, and then contacted with a cobalt-molybdenum catalyst in ebullating-bed reactors. The processing downstream of the dissolver is similar to that used for the SRC II - H_2 gas process; see Figure A-3.

The ebullating-bed reactors are similar, in principal, to those used in commercial H-Oil units. In both H-Coal and H-Oil, the catalyst is suspended in upflowing fluid: In H-Coal, the feed is a slurry; whereas, in H-Oil, it is a heavy oil. This design is ideally suited to the handling of coal-oil slurry, because both the coal particles and the catalyst will be fluidized. The ebullating bed makes it possible to continuously add and withdraw the catalyst and to continuously withdraw unconverted coal solids. The catalyst particles are much larger and they remain in the reactor, while unconverted coal is entrained in the liquefied coal overflow stream. Because the hydroliquefaction step is catalytic, it produces a relatively large amount of light liquids such as naphtha.

The flow rates shown on Figure A-5 were calculated by computer-assisted process simulation.

Heat and Material Balance Factors

Raw water usage	21,500 gal/min
ROM coal charged	72,300 TPSD
Washed & dried coal charged	50,913 TPSD
Heating value of fuel products	832 MMM Btu/SD HHV

Feeds

	TPD	MMM Btu/SD HHV
Coal to dissolver	25,000	630
Coal to gasifier	15,450	390
Coal to low Btu gasifier	10,470	260
Total	50,920	1,280

Products

	TPD	MMM Btu/SD HHV
Pipeline gas	4,930	220
LPG	1,120	50
Naphtha	3,620	160
SRC (400°) 0.71 wt % S	11,100	400
Sulfur	1,620	—
Total	22,390	830

Indicated thermal efficiency, 65%

Estimated Operating Labor

The estimated operating labor personnel is 66 operators per shift.

SYNTHOIL

Data Sources and Status

Data sources included a review of the available process reports prepared by the Pittsburgh Energy Research Center (PERC) funded by the U.S. Bureau of Mines and ERDA, plus a visit to the PERC experimental facilities and discussions with PERC personnel.

The material balance was based on the results of a Synthoil experimental run, No. FB-30, which used Kentucky coal of the following composition:

Proximate Analysis	Wt %	Ultimate Analysis	Wt % MAF Basis
Moisture, wet basis	2.9	Sulfur	5.7
Ash, dry basis	17.4	Oxygen	7.6
Volatile matter, MAF basis	45.5	Carbon	79.1
Fixed carbon, MAF basis	54.5	Hydrogen	6.0
HHV = 12,536 Btu/lb		Nitrogen	1.6

The yields based on the above information were converted to yields expected from Illinois No. 6 coal by elemental balance adjustment as previously described in the subsection on H-Coal.

Process work on the Synthoil process was initiated in 1969 by the U.S. Bureau of Mines. The work is now managed by ERDA through PERC. The first unit processed 5 pounds per hour of coal through a 5/16-inch diameter by 68-foot long reactor. Extended, successful runs of 30 days were made with five different coals.

A larger 1/2-ton-per-day coal unit has been operated for many 500-hour runs. This unit has a 1-inch-diameter by 14-foot-long reactor. A 10-ton-per-hour process development unit is now under construction at PERC.

Process Description

The process and material balance is shown in the Figure A-6 block flow diagram. The process is catalytic and converts low-quality, high-sulfur coals to nonpolluting utility fuel oil. The novel feature of the process is a highly turbulent, concurrent, up-flow, packed-bed reactor in which coal is reacted with hydrogen in the presence of Co-Mo/SiO_2-Al_2O_3 catalyst. At reaction temperatures and pressures of 800° to 850°F and 2,000 to 4,000 psig, respectively, coal is converted to a heavy liquid hydrocarbon; the sulfur is largely converted to H_2S. The process has been demonstrated to be applicable to a wide variety of U.S. coals. Results of Synthoil process studies using a high sulfur, high ash coal feed indicate the high yields of an oil containing 0.2 wt % sulfur.

Figure A-6 shows powdered coal, which is slurried in a portion of the product oil, being combined with a mixture of recycle and fresh hydrogen, heated, and being introduced into the bottom of the reactor. The reactor product stream enters a disengaging drum, and the liquid phase is fed to a filtration section where solids are separated from liquids. The downstream processing is similar to the SRC II - H_2 gas mode; see Figure A-3.

Heat and Material Balance Factors

Raw water usage	19,800 gal/min
ROM coal charged	65,600 TPSD
Washed & dried coal charged	46,200 TPSD
Heating value of fuel products	730 MMM Btu/SD HHV

Feeds

	TPD	MMM Btu/SD HHV
Coal to dissolver	25,000	630
Coal to gasifier	15,320	380
Coal to low Btu gasifier	5,880	150
Total	46,200	1,160

Products

	TPD	MMM Btu/SD HHV
Pipeline gas	5,450	250
LPG	None	—
Naphtha	2,400	100
400°+ fuel oil (0.33% S)	11,420	380
Sulfur	1,510	—
Total	20,780	730

Indicated thermal efficiency, 63%

Estimated Operating Labor

Operating labor is estimated to be 66 operators per shift.

CSF DONOR SOLVENT PROCESS

Data Sources and Status

This analysis is based upon data, preliminary design work, and FCI estimates presented in the OCR R&D Report No. 70, titled "Engineering Evaluation and Review of Consol Synthetic Fuel Process," prepared under Contract 14-32-0001-1217. The 1971 capital estimate appearing in this report was escalated to 1975 for use in this study.

The OCR R&D Report No. 70 was based on the use of West Virginia coal with the following composition:

Proximate Analysis	Wt %	Ultimate Analysis	Wt % MAF Basis
Moisture, wet basis	14.4	Carbon	79.5
Ash, dry basis	13.2	Hydrogen	5.6
Volatile matter, MAF basis	45.0	Nitrogen	1.5
Fixed carbon, MAF basis	55.0	Sulfur	4.9
HHV = 10,820 Btu/lb		Oxygen	8.5

Background information also included the progress reports that described work done in 1966 to 1970 at the Cresap, West Virginia, pilot plant and published under OCR Contract No. 14-01-0001-310. Also used as background information was the conceptual design and economic evaluation published by Parsons as OCR R&D Report No. 45, Interim Report No. 2, titled "1969 Feasibility Report/Consol Synthetic Fuel Process/Snythetic Crude Production"; this work was done under OCR Contract No. 14-01-0001-225.

A 20-TPD coal pilot plant was operated for 33 months from May, 1967, to April, 1970, and operations were then suspended. The facility processed 1,400 tons of coal. Over 600 tons of coal were converted at selected process conditions in the final 3 months of scheduled operations. Mechanical and process problems were defined during the course of operations. At the end of the operating period, uninterrupted runs of the extract production operation of up to 10 days duration were completed and terminated voluntarily.

The pilot plant is presently being reactivated, and startup is expected in 1977.

Process Description

The process is shown schematically in Figure A-7. Here, the coal is extracted in the extractor by a donor solvent derived from the coal. The hydrogenated donor solvent serves as a hydrogen carrier. After extraction, the solids are removed by a combination of partial separation in hydroclones and low-temperature carbonization. The low-solids extract is then fractionated, and a portion is hydrogenated in an ebullating, catalytic bed to produce a wide range of gas and liquid fuels. The hydrotreated product is fractionated and a portion of the liquid is recycled to the extractor as donor solvent. Hydrogen is generated by steam-oxygen gasification of the unextracted char. Clean fuels, hydro residue, and some of the gas produced in the plant are used as plant fuel. Sulfur and ammonia are recovered as byproducts. The product gas and liquid products are available for sale.

This process configuration differs from the SRC, H-Coal, and Synthoil processes previously described; differences include the use of West Virginia feed coal and the use of hydroclones, plus a low-temperature carbonization process instead of filtration and filter cake drying. However, this evaluation incorporated a filtration of the liquid product for solids removal. The fuel oil product has an estimated 0.3% sulfur content.

The OCR R&D Report No. 70 conceptual design was modified by the addition of an air-blown gasifier for captive fuel requirements and a power plant to eliminate the use of outside power.

Heat and Material Balance Factors

The following factors are based on a coal feed of 25,000 TPD West Virginia coal to extraction.

Raw water usage	11,900 gal/min
ROM coal charged	41,400 TPSD
Washed & dried coal charged (MF)	28,350 TPSD
Heating value of fuel products	544 MMM Btu/SD HHV

Feeds

	TPD	MMM Btu/SD HHV
Dried coal to extraction section	25,000	620
Dried coal to low Btu gasifier	3,350	90
Total	28,350	710

Products

	TPD	MMM Btu/SD HHV
Pipeline gas	2,260	90
Naphtha	2,000	75
Fuel oil (0.30 wt % S)	10,360	375
Sulfur	1,190	—
Total	15,810	540

Indicated thermal efficiency, 76%

Estimated Operating Labor

The operating labor is estimated to be 66 operators per shift.

COED

Data Sources and Status

The principal data source was ERDA R&D Report No. 114 - Interim Report No. 1, titled "Commercial Complex, Conceptual Design/Economic Analysis, Oil and Power by COED Based Coal Conversion" (1975). This report was developed under ERDA Contract No. E(49-18)-1775.

The design used in this study differs from the R&D Report No. 114 design in that the in-process char is converted to 1,000 Btu SNG; the design in the ERDA R&D Report No. 114 produced electrical power from this gas.

This process has been developed through the pilot plant stage by the FMC Corporation at Princeton, New Jersey. Over 20,000 tons, including seven different coals, were processed in the 36-TPD pilot plant. The pilot plant was operational from 1970 to 1974, achieved its objectives, and was then shut down.

The original work began in 1956, and OCR sponsorship began in 1962.

At present, plans are under way for a demonstration plant, under ERDA sponsorship, using the COED fluidized-bed pyrolysis process. Planned products are SNG and liquid fuels.

Process Description

The process and material balance are shown in the block flow diagram, Figure A-8. It produces 28,000 BPD of 25° API, 0.1 wt % sulfur syncrude, M SCFD of 1,000 Btu SNG, and 766 LTSD of sulfur from 35,700 TPD ROM Illinois No. 6 coal.

The process consists of pyrolyzing feed coal in fluidized-bed reactors; the pyrolyzed volatiles, containing high Btu gas and raw COED oil (a tar), are separated. The H_2S in the gas is removed by absorption to reduce high Btu fuel gas, and the tar is filtered to remove solids and then hydrotreated to produce syncrude. The heat required to sustain the pyrolysis process is supplied by gasifying char produced in the pyrolysis reactors; the gasification is done in fluid-bed units using oxygen and steam in a reducing atmosphere. Part of the syngas produced by char gasification is used for hydrogen production and, after H_2S removal, for captive fuel gas in the plant and low Btu product fuel gas. High and low Btu fuel gases are combined to make plant fuel gas for power and steam generation, plus product pipeline gas. The estimated thermal efficiency for this COED configuration is 55%, including byproduct sulfur.

Heat and Material Balance Factors

Raw water usage	28,800 gal/min
ROM coal charged	35,700 TPSD
Washed & dried coal charged	25,000 TPSD
Heating value of fuel products	354,340 MMM Btu/SD HHV

Feed

	TPSD	MMM Btu/SD HHV
Washed & dried coal	25,000	610

Products

	TPSD	MMM Btu/SD HHV
Fuel gas (300 Btu/SCF, 250 psig)	3,580	160
Syncrude COED oil (0.1% sulfur, 25° API)	4,430	170
Sulfur	680	5
Total	8,690	335

Indicated thermal efficiency, 55%

Operating Manpower

The estimated manpower requirement is 49 operators per shift.

COALCON

Data Sources and Status

Technical and economic information for this study originated primarily in the report titled "Conceptual Commercial Design and Commercial Feasibility Evaluation for Clean Boiler Fuel Facility," prepared under ERDA Contract E(49-18)-1772 (Ref. 6) by the Dravo Corporation.

The reference design and economics are based on processing 50,100 TPD of feed coal to produce SNG, LPG, fuel oils, ammonia, sulfur, pyridine, and phenol. This conceptual design capacity was reduced to process 25,000 TPSD through the prime coal conversion steps using appropriate scaling factors. Design yields for the coal conversion step, a fluid-bed hydropyrolysis, were taken directly from a previously published design by Coalcon, the process developer (Ref. 5).

An extensive pilot plant program was carried out in the early 1960s by the Union Carbide Corp., using a low-sulfur, Western subbituminous coal (Lake DeSmet). The process was tested on three scales: a 1-pound-per-hour bunch scale reactor, a 10-pound-per-hour small pilot, and a 1-ton-per-hour pilot plant. The design of a 2,600-TPD demonstration plant was begun in 1975 under ERDA sponsorship.

Process Description

The process block flow diagram and material balance are shown in Figure A-9. The heart of the process is a pressurized, fluid-bed hydropyrolyzer, in which coal is reacted to produce partially desulfurized volatile products and char. Reaction conditions are about 1,050°F and 550 psig. The hydrogen consumption is approximately 31,300 SCF per ton of coal feed. Products of hydrocarbonization include gases, water soluble chemicals (phenols and pyridine), tar, and char. Gases are treated to recover ammonia and remove acid gases, and they are then sent to a cryogenic separation plant to produce recycle hydrogen, SNG, and LPGs. Oils are fractionated into light and medium fuel oil products and a heavy oil that is used as plant fuel for steam generation. Char is gasified with steam and oxygen to a carbon monoxide and a hydrogen-rich gas for process use.

Heat and Material Balance Factors

The following factors for the Dravo Coalcon plant are based on a coal feed of 25,000 TPD Pittsburgh No. 8 seam coal.

Raw water usage	21,400 gal/min
Electric power usage	75 MW (18 MM Btu/day equivalent heat)
ROM coal charged	37,500 TPSD
Washed & dried coal charged	25,000/SD

Feeds

	TPD	MMM Btu/SD HHV
Dried coal	25,000	590
Electrical power (150 MM)	–	20
Total	25,000	610

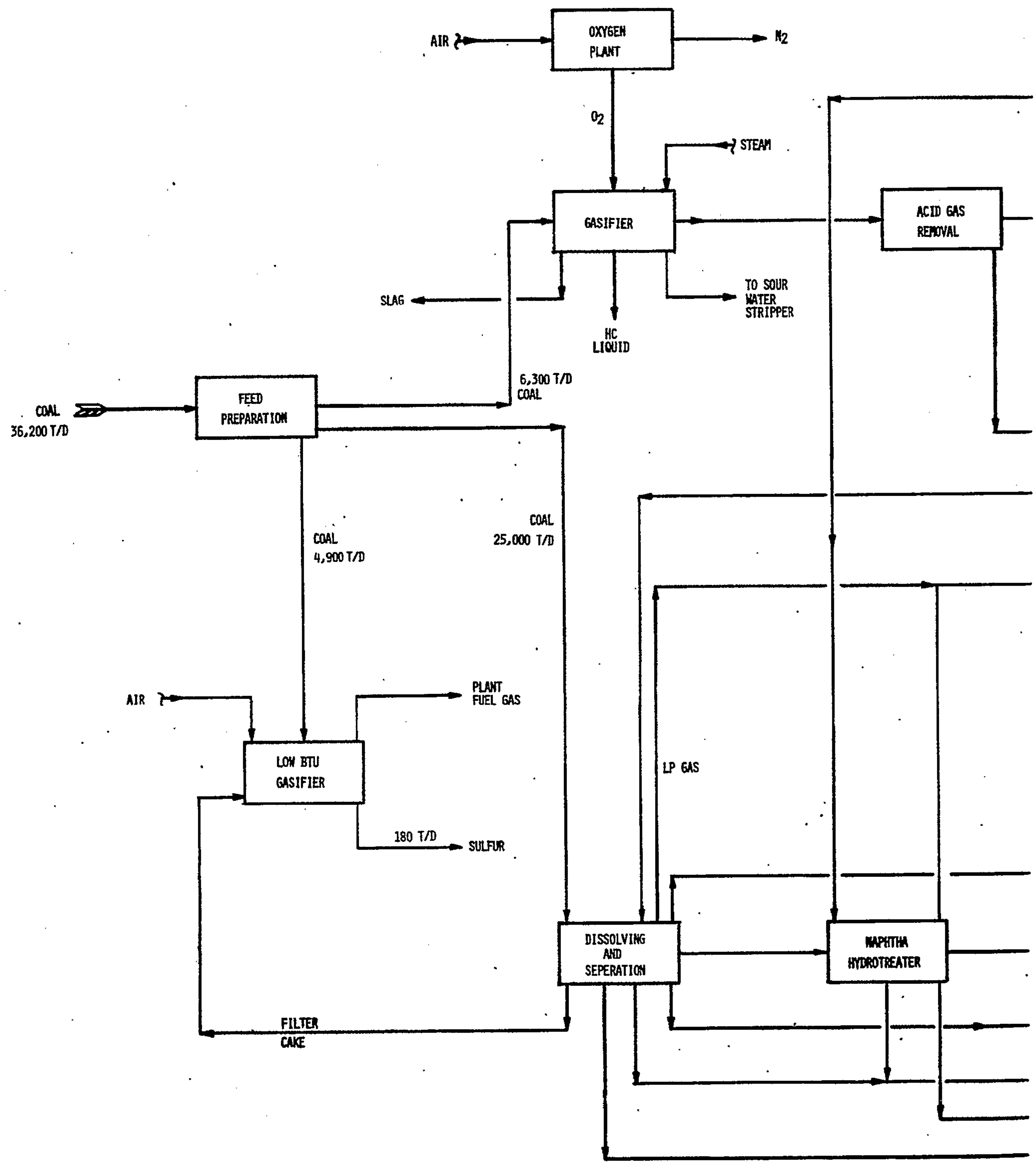
Products

	TPD	MMM Btu/SD HHV
SNG	3,460	160
LPG	595	25
Light oil	1,075	40
Fuel oil	3,725	130
Phenols	100	3
Pyridine	870	15
Sulfur	535	4
Ammonia	100	2
Total	10,460	379

Indicated thermal efficiency, 62%

Estimated Operating Labor

The operating labor is estimated by Dravo to be 130 operators per shift.



AIR
OXYGEN PLANT
N_2
O_2
STEAM
GASIFIER
ACID GAS REMOVAL
SLAG
HC LIQUID
TO SOUR WATER STRIPPER
6,300 T/D COAL
COAL 36,200 T/D
FEED PREPARATION
COAL 4,900 T/D
COAL 25,000 T/D
AIR
PLANT FUEL GAS
LOW BTU GASIFIER
180 T/D
SULFUR
LP GAS
DISSOLVING AND SEPERATION
NAPHTHA HYDROTREATER
FILTER CAKE

Fi

A

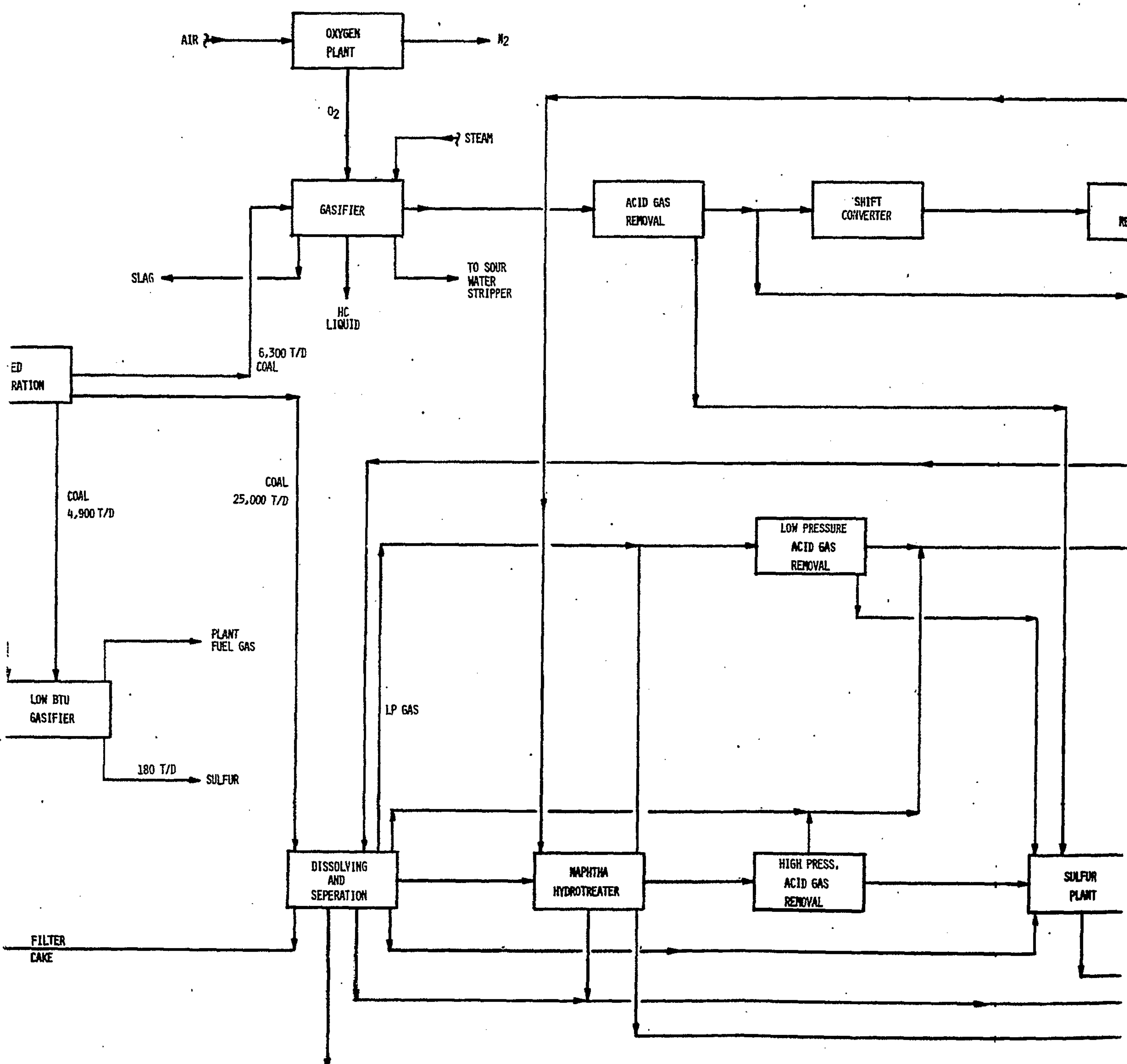
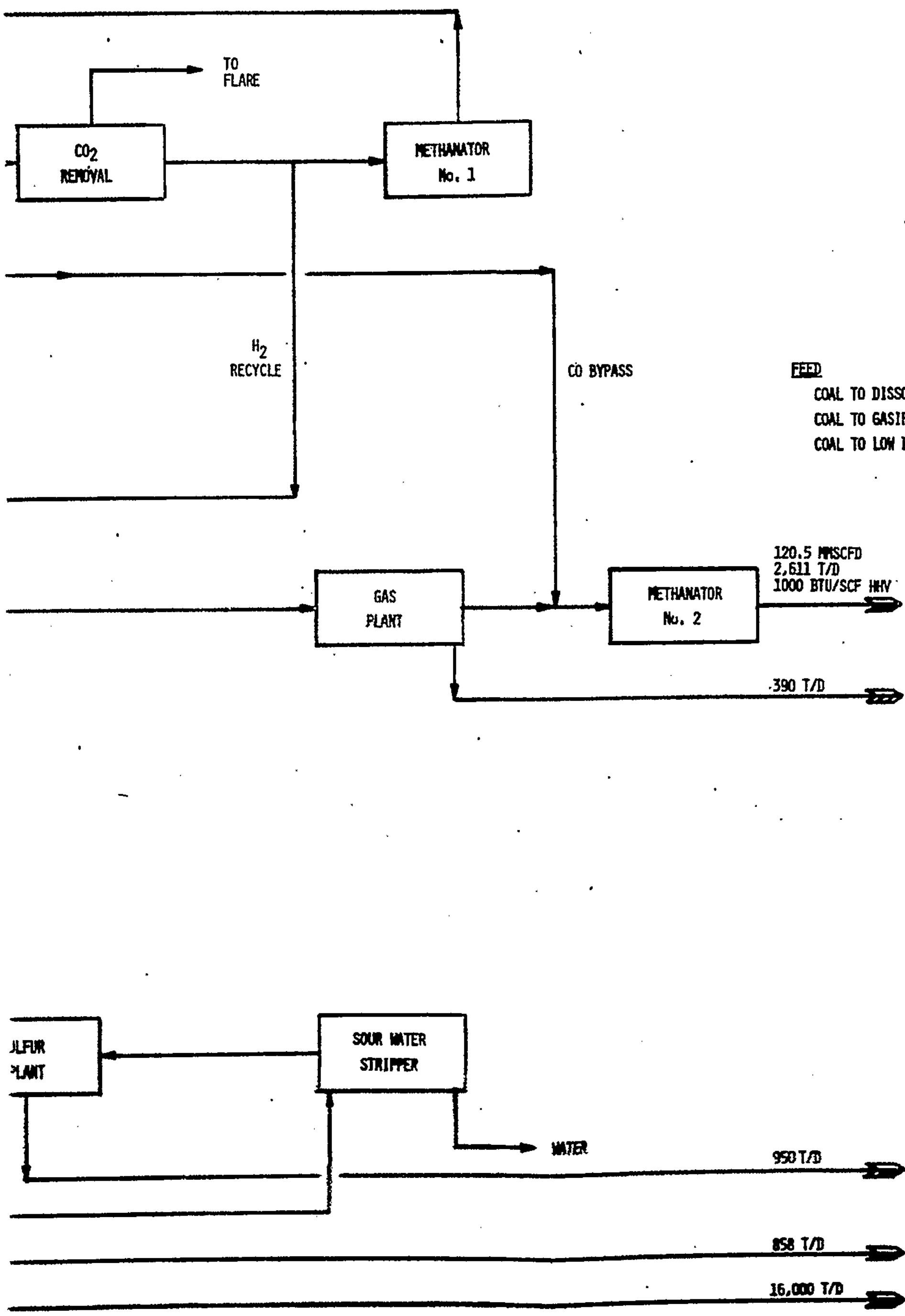


Figure A-1 POGO Plant Block Flow Diagram, SRC I - H

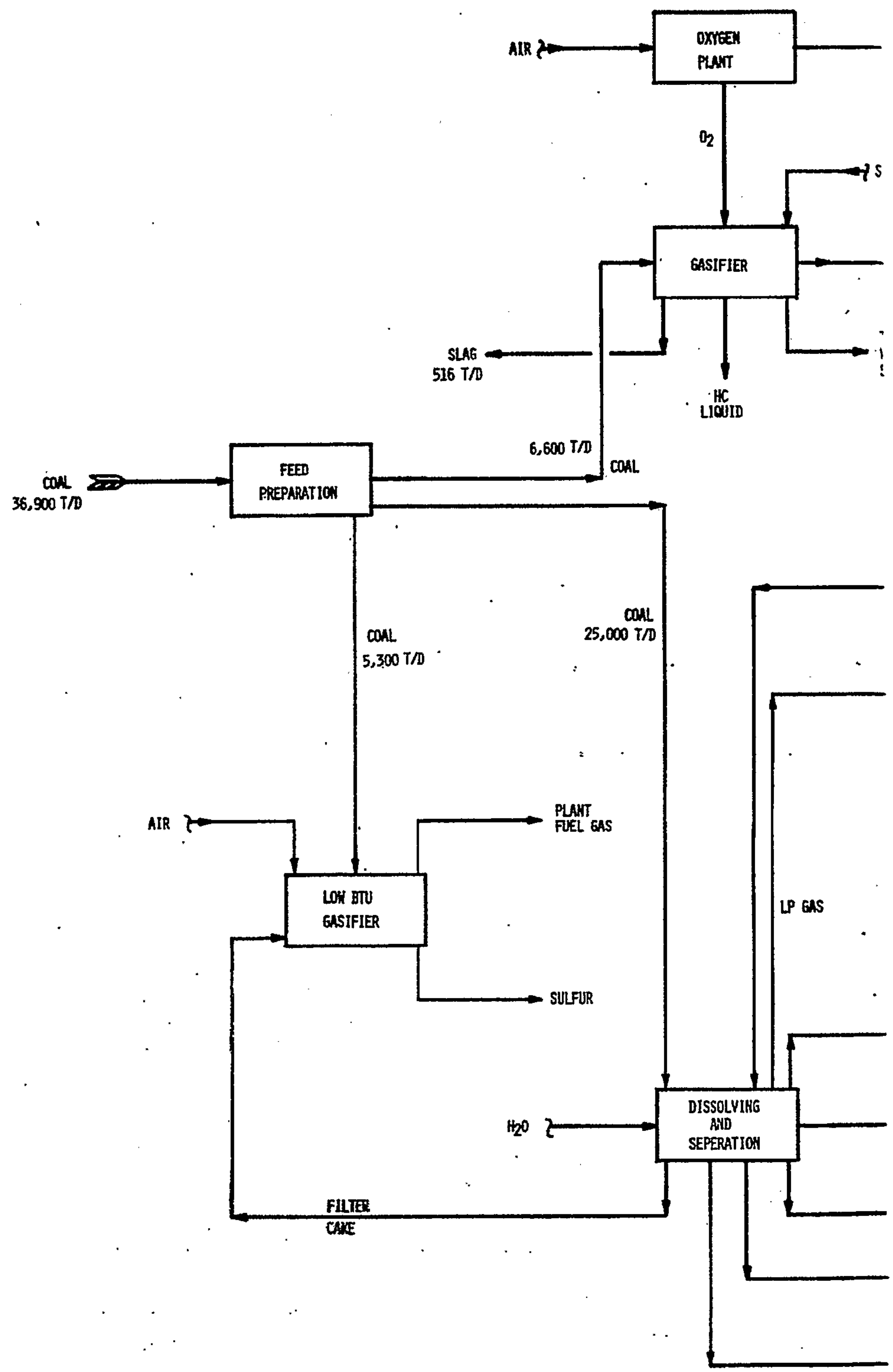
B



	T/D	HHV MMBTU/DAY
FEED		
COAL TO DISSOLVER	25,000	630
COAL TO GASIFIER	6,300	160
COAL TO LOW BTU GASIFIER	4,900	120
TOTAL	36,200	910
PIPELINE GAS	2,600	120
LPG	390	20
SULFUR	1,130	
NAPHTHA	860	40
400 + PRODUCT	16,000	500
TOTAL		680

C I – H_2 Gas

C



AIR
OXYGEN PLANT
O_2
GASIFIER
SLAG 516 T/D
HC LIQUID
6,600 T/D
COAL
COAL 36,900 T/D
FEED PREPARATION
COAL 5,300 T/D
COAL 25,000 T/D
AIR
PLANT FUEL GAS
LOW BTU GASIFIER
SULFUR
LP GAS
H_2O
DISSOLVING AND SEPERATION
FILTER CAKE

A

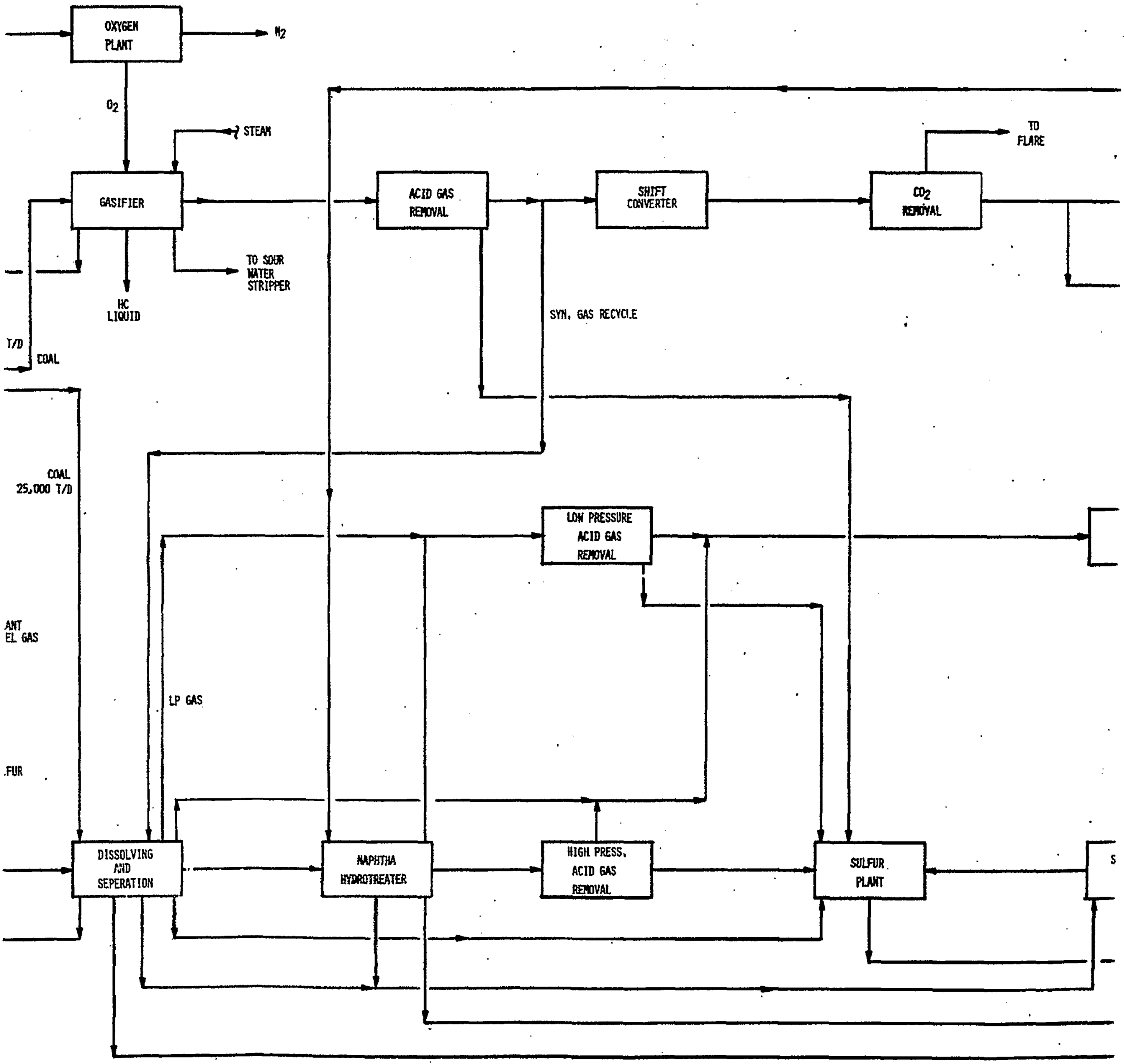
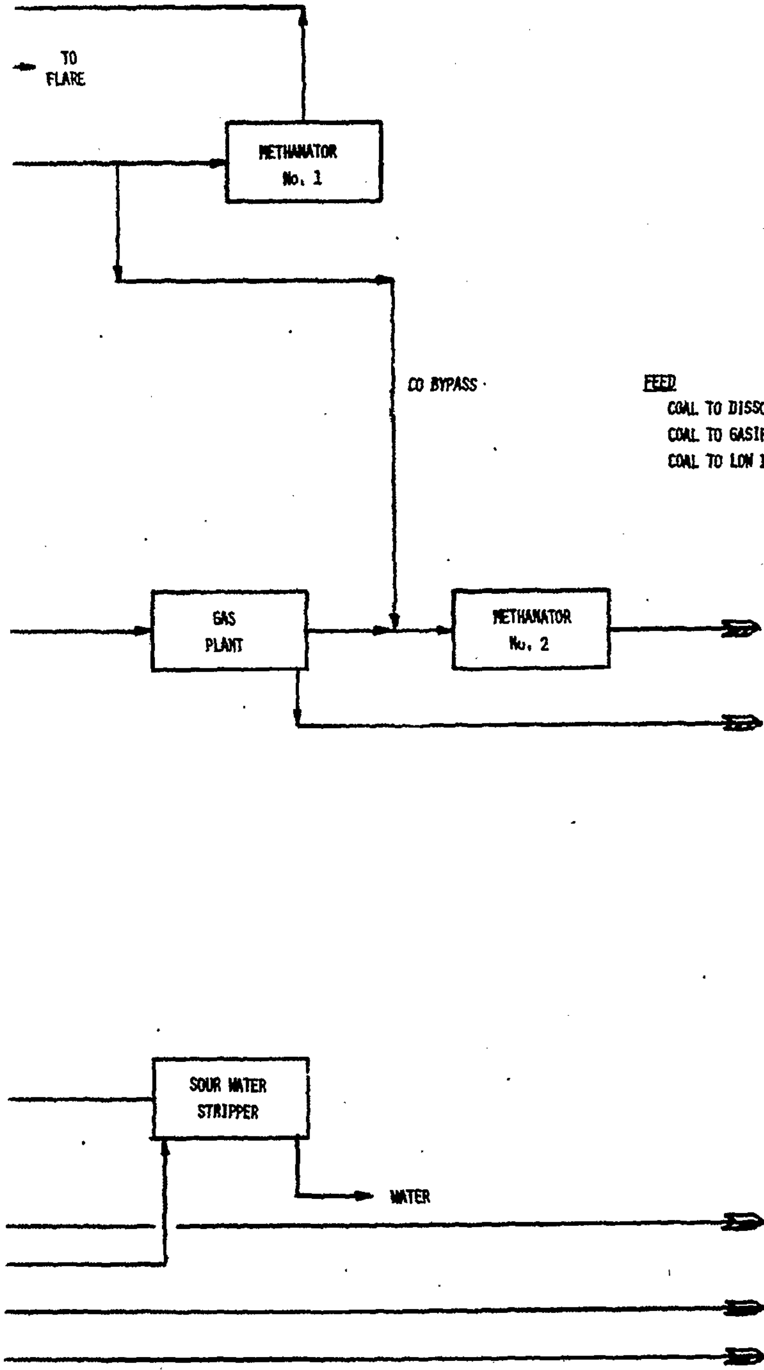
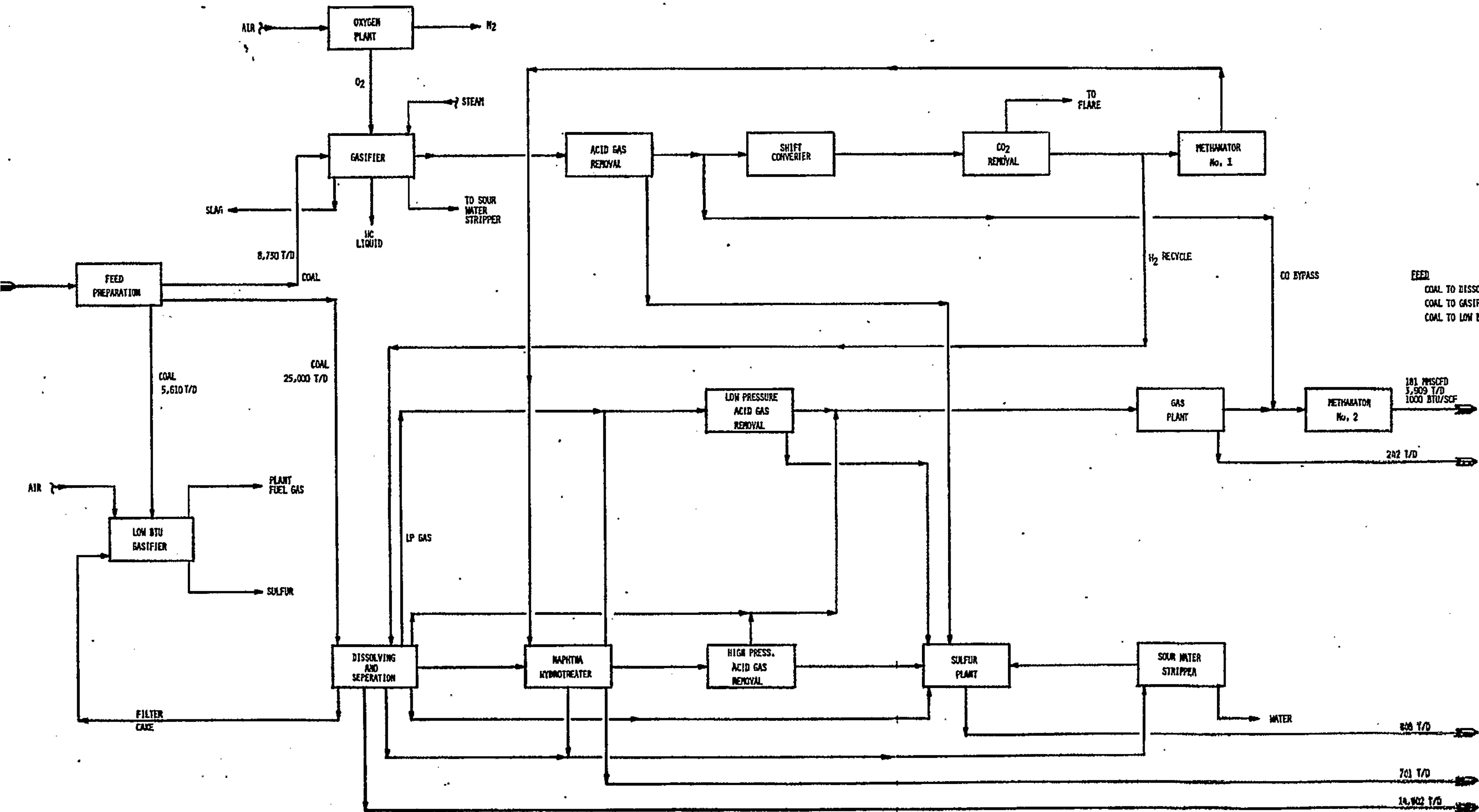


Figure A-2 POGO Plant Block Flow Diagram, SRC I - Syngas



	T/D	HHV MMMBTU/DAY
FEED		
COAL TO DISSOLVER	25,000	630
COAL TO GASIFIER	6,600	170
COAL TO LOW BTU GASIFIER	5,300	130
TOTAL	36,900	930
PIPELINE GAS	2,800	130
LPG	240	10
SULFUR	1,150	
NAPHTHA	860	40
400+ PRODUCT	16,000	500
TOTAL		680



	T/D	HHV MMBTU/DAY
FEED		
COAL TO DISSOLVER	25,000	630
COAL TO GASIFIER	8,730	220
COAL TO LOW BTU GASIFIER	5,610	140
TOTAL	39,340	990
PIPELINE GAS	3,910	180
LPG	240	10
SULFUR	1,240	
NAPHTHA	700	30
400+ PRODUCT	14,900	500
TOTAL		720

Figure A-3 POGO Plant Block Flow Diagram, SRC II - H_2 Gas

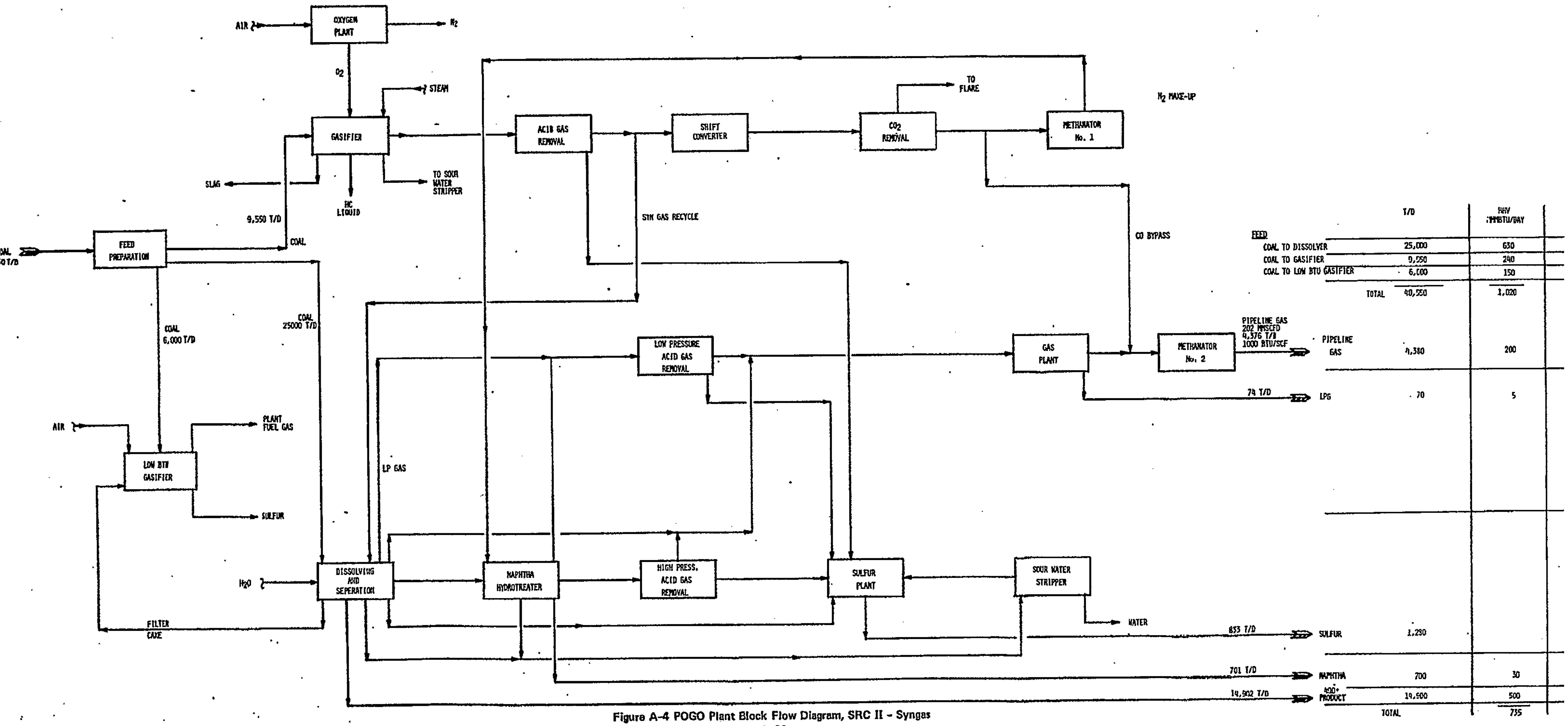


Figure A-4 POGO Plant Block Flow Diagram, SRC II - Syngas

B

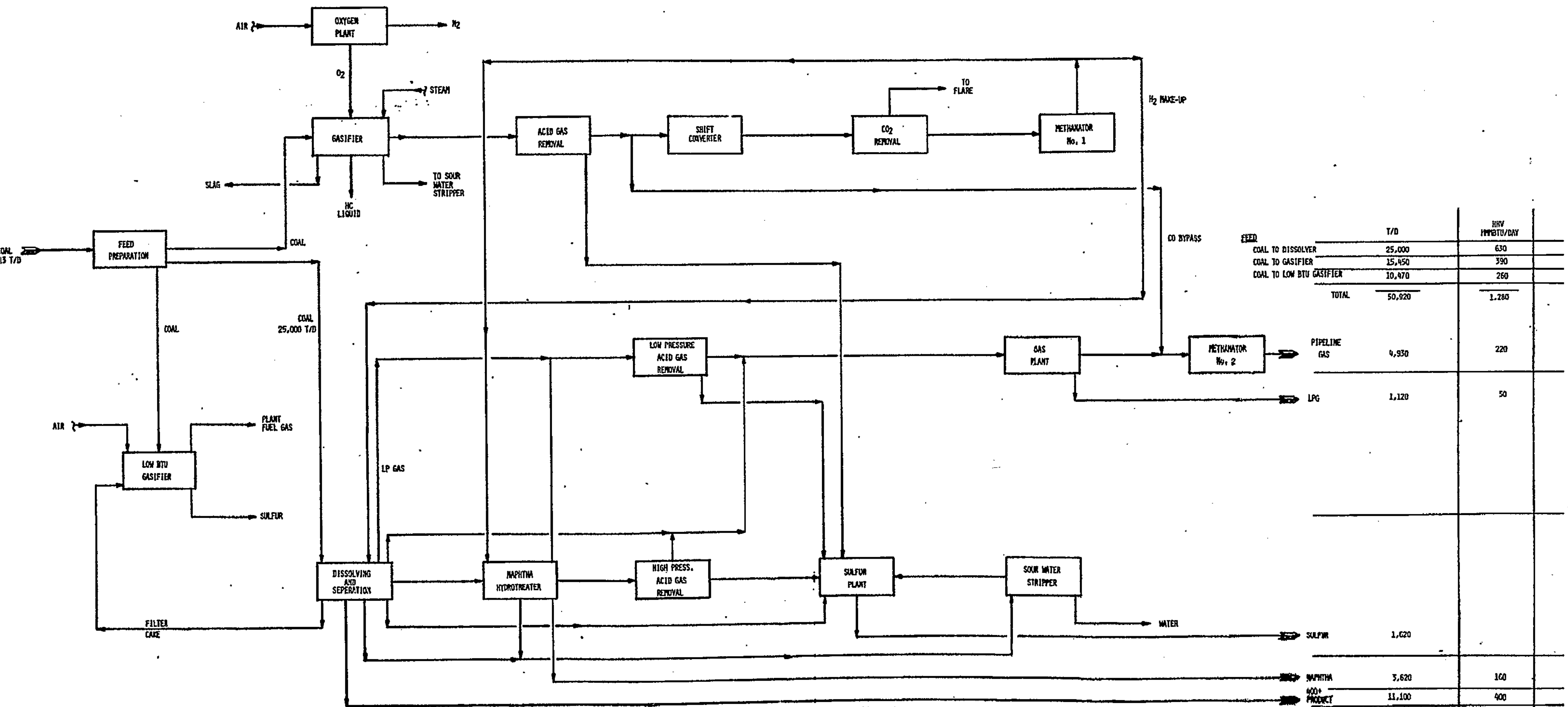
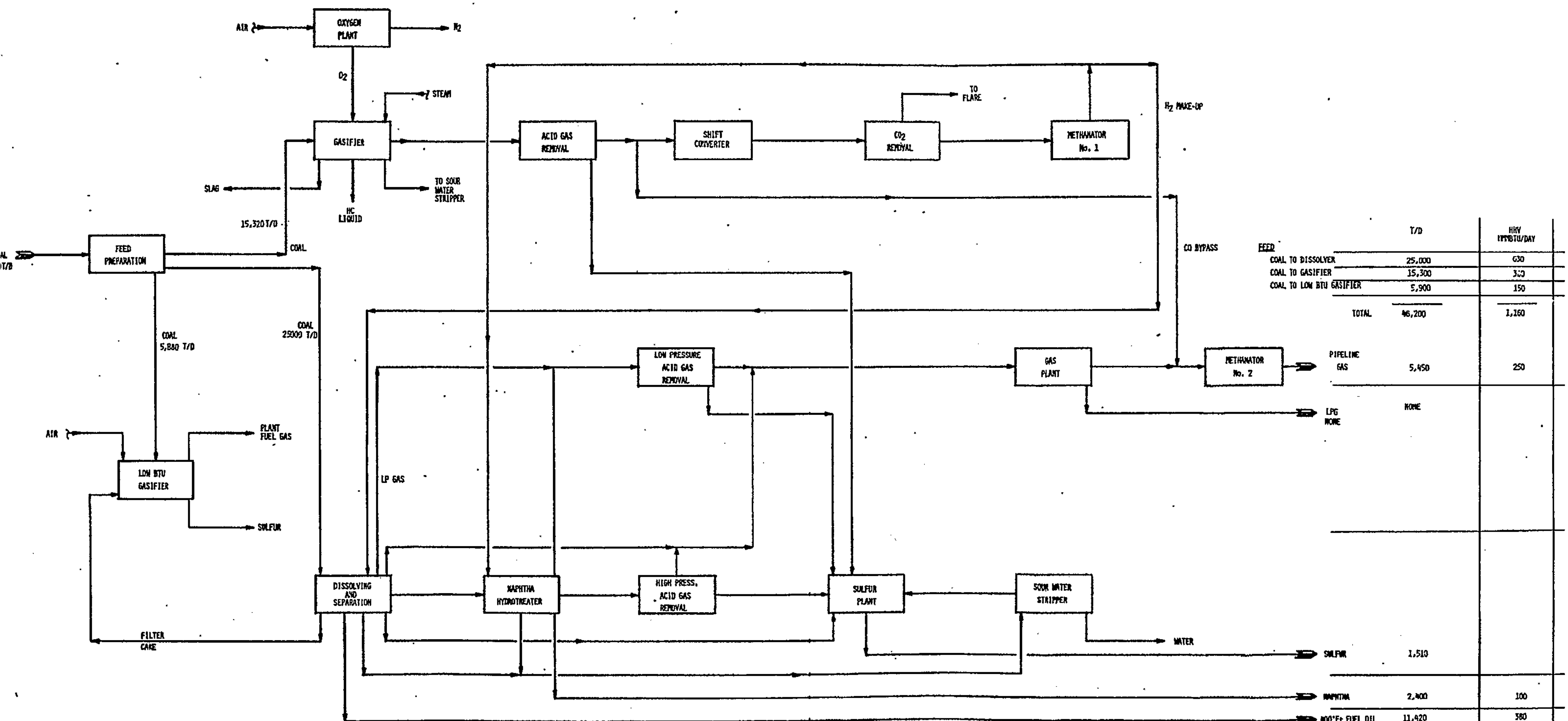


Figure A-5 POGO Plant Block Flow Diagram, H-Coal



	T/D	HHV MMBTU/DAY
FEED		
COAL TO DISSOLVER	25,000	630
COAL TO GASIFIER	15,300	3?0
COAL TO LOW BTU GASIFIER	5,900	150
TOTAL	46,200	1,160
PIPELINE GAS	5,450	250
LPG	NONE	
SULFUR	1,510	
NAPHTHA	2,400	100
400°F+ FUEL OIL	11,420	380
TOTAL		730

Figure A-6 POGO Plant Block Flow Diagram, Synthoil

13

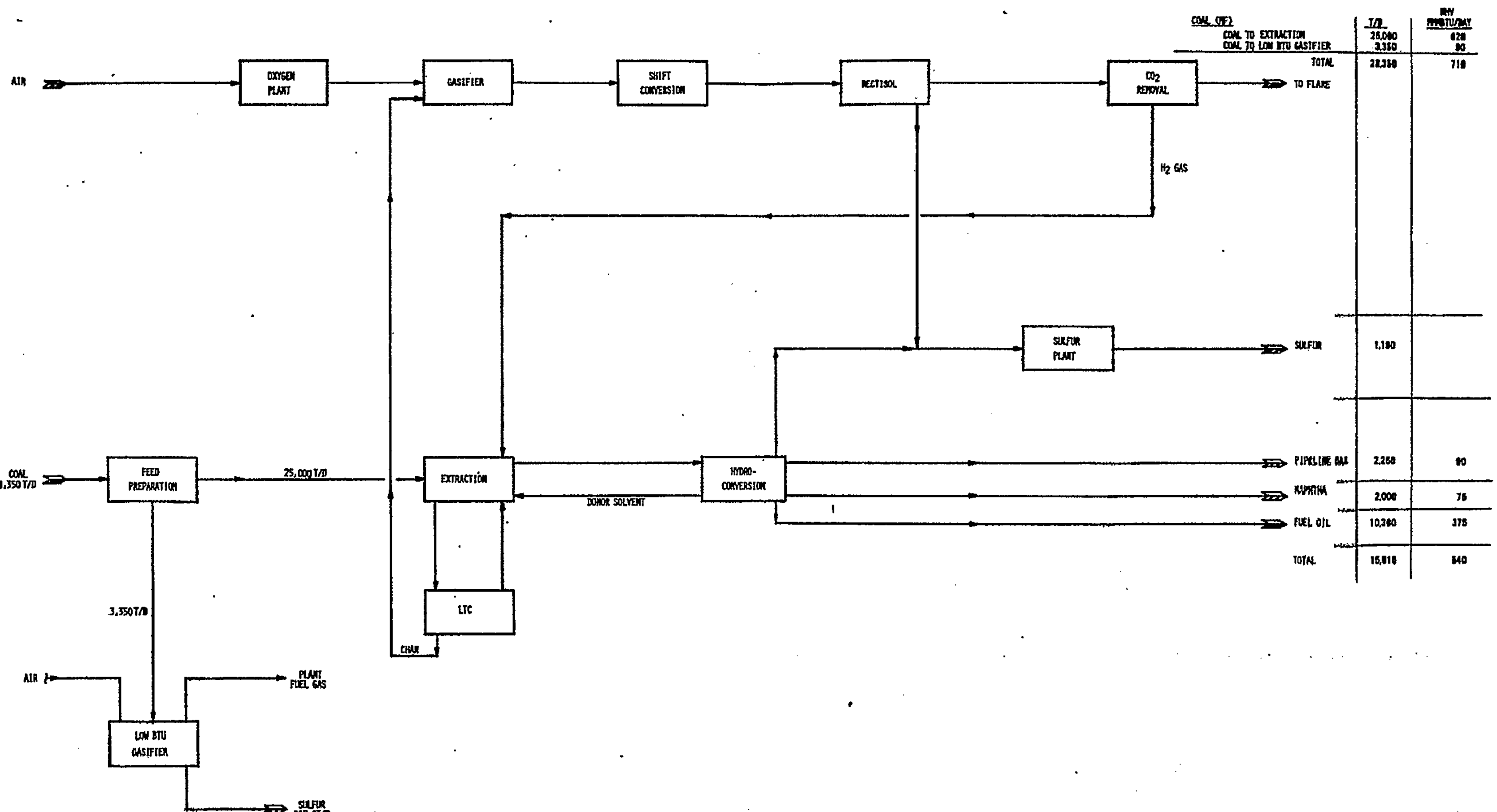


Figure A-7 POGO Plant Block Flow Diagram, CSF Doner Solvent

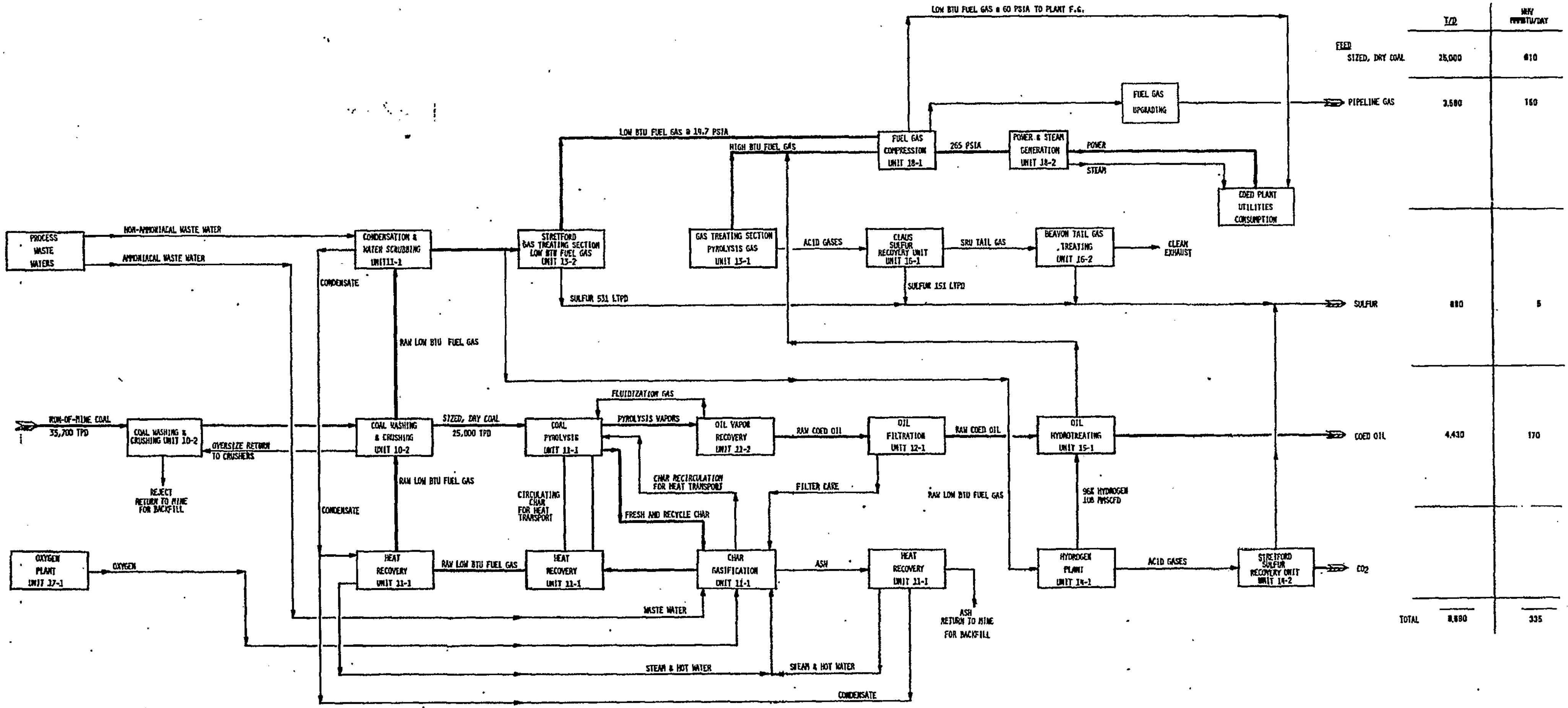


Figure A-8 POGO Plant Block Flow Diagram, COED

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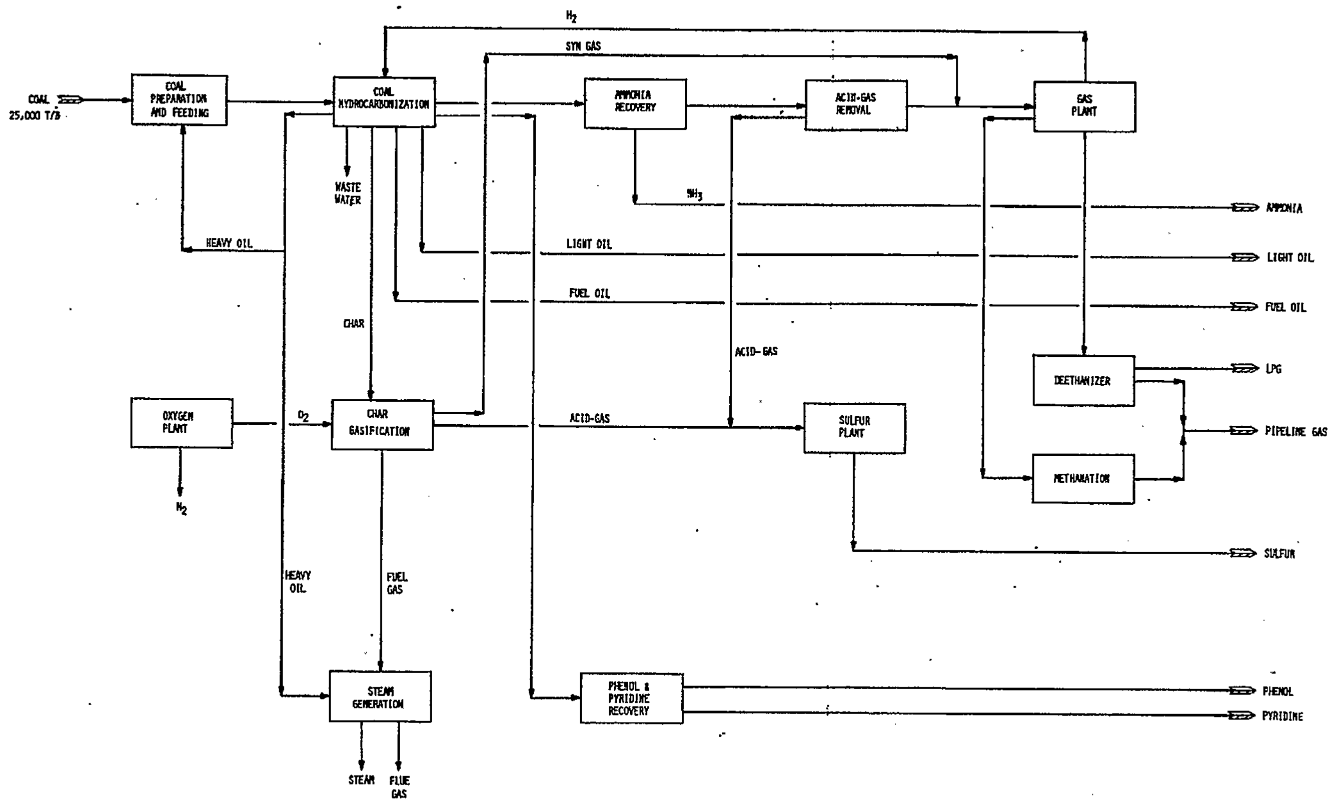


FEED	T/D	HHV MMBTU/DAY
COAL	25,000	590
POWER		20
TOTAL		610
	100	2
	1,075	40
	3,725	130
	595	25
	3,480	160
	535	4
	100	3
	870	15
TOTAL	10,480	378

Figure A-9 POGO Plant Block Flow Diagram, Coalcon Process

APPENDIX B

FISCHER-TROPSCH AND OIL/GAS

Process descriptions and characteristics of the then in-progress Fischer-Tropsch and Oil/Gas designs are presented. Revisions and improvements were made to these designs prior to publication of their final ERDA R&D reports (Refs. 7 and 8).

FISCHER-TROPSCH PLANT

The Fischer-Tropsch is a conceptual design based in part on in-progress experimental results of flame-sprayed catalyst systems obtained by the ERDA Pittsburgh Energy Research Center (PERC). The design is intended to illustrate the potentials for a large, second-generation complex practicing Fischer-Tropsch technology. Its reduction to practice will depend on the successful completion of a number of in-progress development programs, plus pilot plant work that has not yet started.

Process Description

The Fischer-Tropsch complex, depicted in Figure B-1, includes a captive coal mine that produces approximately 40,000 TPD of run-of-mine (ROM) coal. This coal is cleaned and sized, and about 30,000 TPD of the prepared Illinois No. 6 coal is consumed each day to produce over 500 billion Btu/day of products. The coal is completely gasified in two, 2-stage, entrained, slagging-type gasifiers. Part of the effluent gas is sent to an isothermal, sour-shift reactor, where CO is shifted with steam to produce the CO/H_2 ratio desired for Fischer-Tropsch reactor feed. The Fischer-Tropsch reactor is a finned-tube heat exchanger with a flame-sprayed iron oxide catalyst on the outside surface. The exothermic reaction is controlled by the generation of high-pressure steam in the tubes while still maintaining a ratio of 1.5:1 cold recycle to fresh feed. The unreacted CO and H_2 are separated from the Fischer-Tropsch liquids and sent to the methanation reactor to produce SNG. The Fischer-Tropsch synthesized liquids are separated by fractionation to produce LPG, light naphtha, heavy naphtha, diesel, and fuel oil products. A chemical recovery section also recovers alcohols.

Heat and Material Balance

The heat and material balances for the Fischer-Tropsch process are summarized in the following table.

Feed	TPD	MM Btu/Hr HHV
Feed coal to gasifier	30,000	31,375
Products		
SNG - 263 MM SCFD	6,600	11,160
LPG - 3,588 BPD	340	600
Naphthas - 20,484 BPD	2,400	4,090
Diesel - 16,318 BPD	2,100	3,550
Fuel oil - 5,033 BPD	715	1,180
Alcohols - 3,971 BPD	460	480
Sulfur	1,015	340
Export power - 139,570 kW @ 33% efficiency	-	1,430
Total	13,630	22,830

Indicated thermal efficiency $\simeq$ 73%

Data Sources and Status

The data used for the Fischer-Tropsch plant is extracted from the conceptual design of a U.S. Fischer-Tropsch plant presently being developed by The Ralph M. Parsons Company under contract to ERDA (Ref. 7).

Preliminary Analysis

The Fischer-Tropsch SNG, at 1,050 Btu/SCF, is an excellent pipeline material. The liquid products contain a high percentage of paraffin compounds; the naphthas have low octane numbers and would preferably be sold as chemical feedstock or reformer feed to a refinery. All products have essentially nil sulfur nitrogen and particulate content.

Further details of this process will be found in the ERDA R&D report (Ref. 7).

OIL/GAS PLANT

The then in-progress Oil/Gas design uses SRC II-type processing with high hydrogenation severity in the dissolver, cryogenic hydrogen separation, and a high-pressure gasifier. It uses preliminary data generated by the ERDA SRC pilot plant located at Fort Lewis, Washington, with an extrapolation to high severity to produce significant gas.

Process Description

The process is depicted in Figure B-2.

The complex includes a captive coal mine to supply approximately 47,000 tons of ROM coal per stream day plus facilities to prepare 35,570 tons per stream day (TPSD) clean, sized coal as feed to the process units.

Facilities to produce oxygen and all required utilities are provided, as are facilities for the treatment and disposal of solid, liquid, and gaseous effluent streams.

Key coal conversion units are:

- A three-train hydroliquefaction unit to convert 20,000 TPSD of feed coal to the primary products: SNG, LPG, naphtha, and fuel oil.
- A process gasifier to convert 10,000 TPSD of feed coal to methane, syngas, and minor amounts of byproducts.
- A fuel-gas gasifier to produce energy for captive use from 5,670 TPSD of coal, plus dry filter cake.

Additional process units shown recover and refine the products plus treat waste streams to produce environmentally acceptable effluents. Further details are contained in the final report (Ref. 8).

Heat and Material Balance

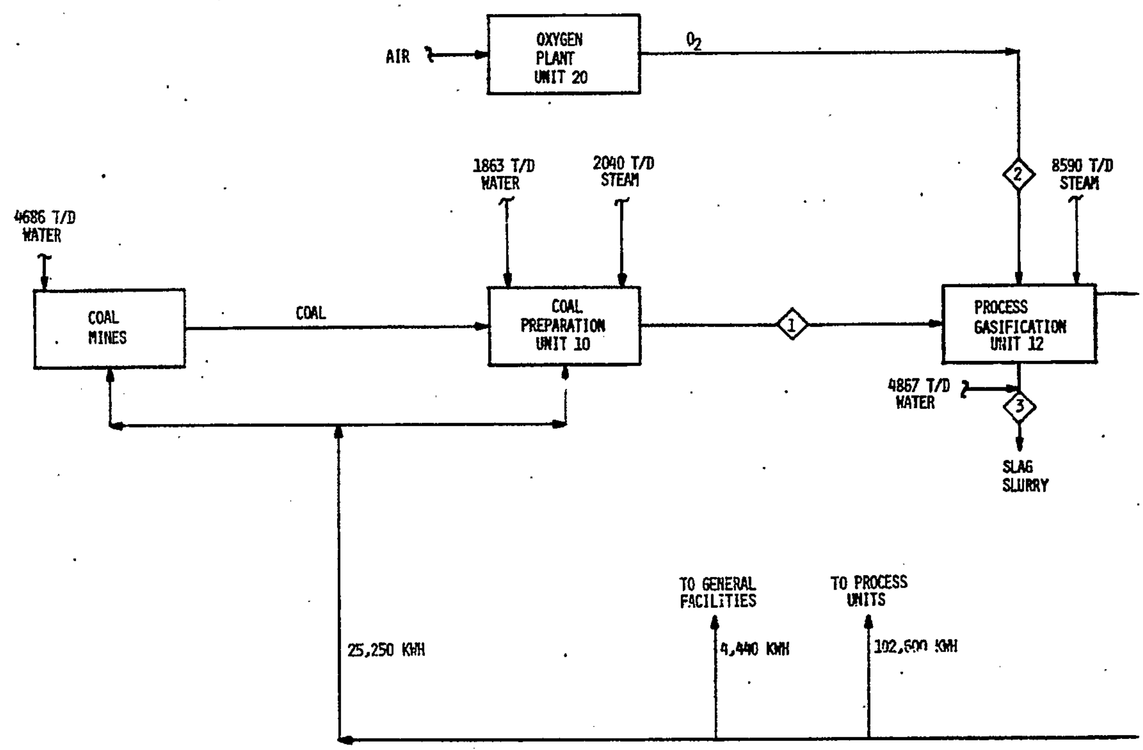
The heat and material balances for the Oil/Gas plant design is summarized in the following table.

Feeds	TPD	MM Btu/Hr HHV
Dried coal to dissolver	20,000	20,200
Dried coal to gasifier	10,000	10,100
Dried coal to low Btu gasifier	5,700	5,700
Total	35,700	36,000

Products	TPD	MM Btu/Hr HHV
SNG	3,900	7,250
C_3 LPG	530	960
C_4 LPG	400	720
Naphtha	1,280	2,260
Fuel oil (400° +) 0.5 wt % S	11,300	16,280
Sulfur	1,250	-
Ammonia	90	-
Total	18,750	27,470

Indicated thermal efficiency $\simeq$ 77%

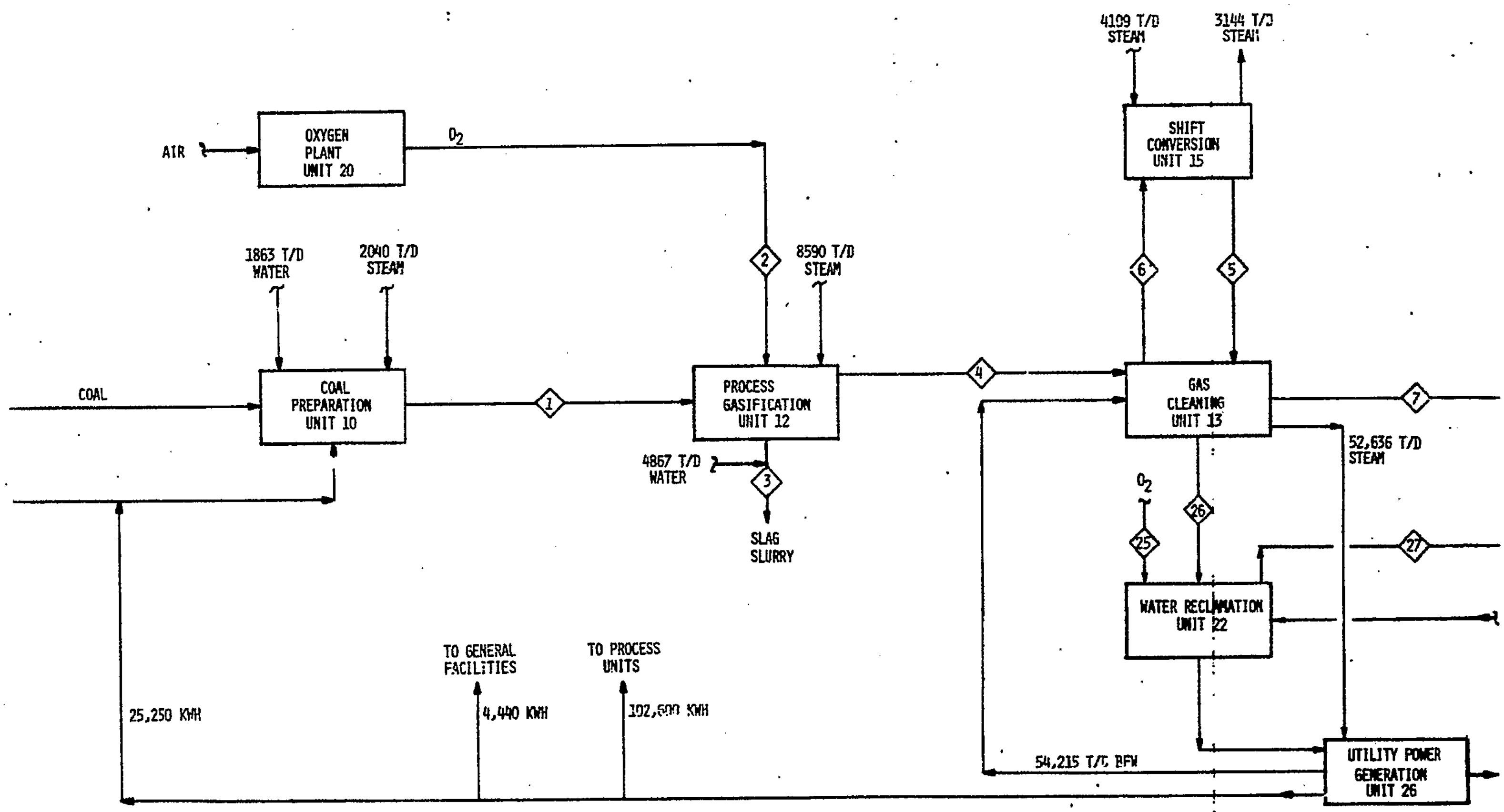
Further details and description of this process will be found in the ERDA R&D report (Ref. 8).



STREAM NO.	1	2	3	4	5	6	7	8	9	10
STREAM NAME	COAL FEED	OXYGEN	SLAG SLURRY	RAW SYN GAS	SHIFTED SYN GAS	SYN GAS TO SHIFT	DUST FREE SYN GAS	CO_2 VENT	ACID GAS	SWEET SYN GAS
Flow, T/D										
H_2	-	-	-	1177	697	361	1513	0.24	-	1512
CO	-	-	-	19172	1213	5886	14499	200 PPMV	-	14376
CO_2	-	-	-	7052	9506	2165	14393	11480	2342	570
CH_4	-	-	-	477	146	146	477	7	-	445
H_2S	-	-	-	526	162	162	519	5 PPMV	519	0.1PPMV
N_2	-	171	-	329	101	101	329	-	8	321
O_2	-	9580	-	-	-	-	-	-	-	-
H_2O	404	-	4886	3409	2152	1047	108	40	28	41
NH_3	-	-	-	4	1.2	1.2	-	-	-	-
COS	-	-	-	19	6	6	19	-	19	-
SO_2	-	-	-	2	0.8	0.8	2	-	2	-
ASH	1070	-	2140	-	-	-	-	-	-	-
CHAR	-	-		351	0.6	0.6	-	-	-	-
COAL (MAF)	13526	-	-	-	-	-	-	-	-	-
TOTAL T/D	15000	9751	7026	32518	13986	9876	31860	11527	2918	17265

STREA
STREA
Flow,
H_2
CO
CO_2
N_2
H_2O
C1
C2-
C2
C3-
C3
C4-
C4
C5-
C5
LIGH
HEAVY
DIESE
FUEL
ALCOH
ACID
TOTAL

A



3 SLAG SLURRY	4 RAW SYN GAS	5 SHIFTED SYN GAS	6 SYN GAS TO SHIFT	7 DUST FREE SYN GAS	8 CO_2 VENT	9 ACID GAS	10 SWEET SYN GAS
-	1177	697	361	1513	0.24	-	1512
-	19172	1213	5886	14499	200 PPMV	-	14376
-	7052	9506	2165	14393	11480	2342	570
-	477	146	146	477	7	-	445
-	526	162	162	519	5 PPMV	519	0.1PPMV
-	329	101	101	329	-	8	321
-	-	-	-	-	-	-	-
4886	3409	2152	1047	108	40	28	41
-	4	1.2	1.2	-	-	-	-
-	19	6	6	19	-	19	-
-	2	0.8	0.8	2	-	2	-
2140	-	-	-	-	-	-	-
	351	0.6	0.6	-	-	-	-
-	-	-	-	-	-	-	-
7026	32518	13986	9876	31860	11527	2918	17265

STREAM NO.	11	12	13	14	15	16
STREAM NAME	ALCOHOL SOLUTION	F-T GAS	F-T LIQUID	C_3/C_4'S	STRIPPED F-T GAS	BUTA[illegible]
Flow, T/D						
H_2	-	397	0.5	-	398	-
CO	-	1632	6	-	1638	-
CO_2	4	138	0.8	13	126	-
N_2	-	328	1	-	329	-
H_2O	2408	27	-	-	-	-
C_1	-	1001	8	4	1005	-
C_2-	-	65	2	6	60	-
C_2	-	258	9	42	226	-
C_3-	-	41	4	26	19	-
C_3	-	141	15	102	53	-
C_4-	-	105	29	106	1	[illegible]
C_4	-	304	98	252	1	7[illegible]
C_5-	-	24	20	1	-	
C_5	-	103	97	-	2	
LIGHT NAPHTHA	-	95	270	-	8	-
HEAVY NAPHTHA	-	47	551	-	2	-
DIESEL	-	7	1103	-	-	-
FUEL OIL	-	-	332	-	-	-
ALCOHOLS	271	-	-	-	-	-
ACID SALTS	41	-	-	-	-	-
TOTAL T/D	2724	4717	2543	552	3869	[illegible]

Figure B-1 Fischer-Tropsch Plant Block Flow Diagram

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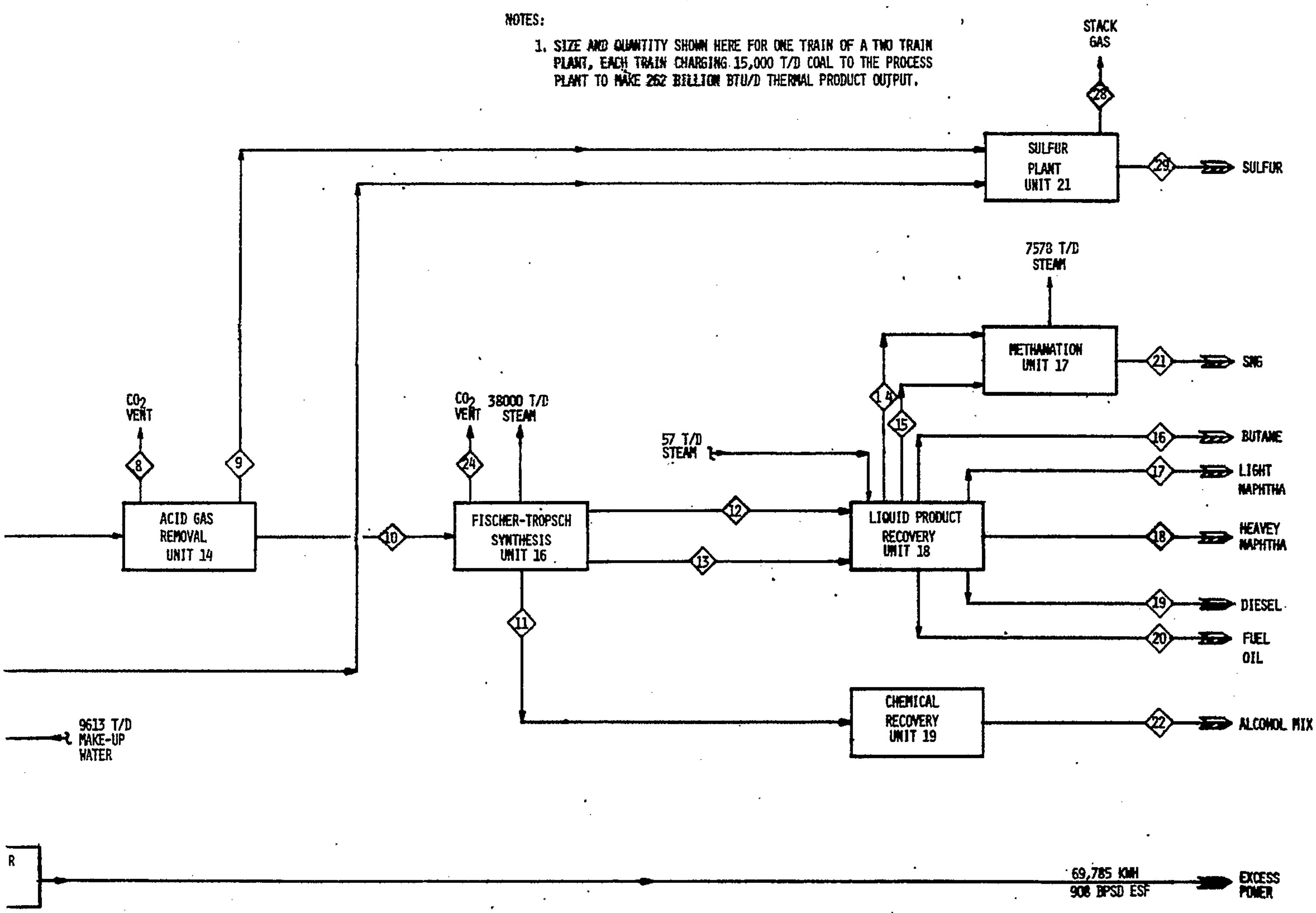
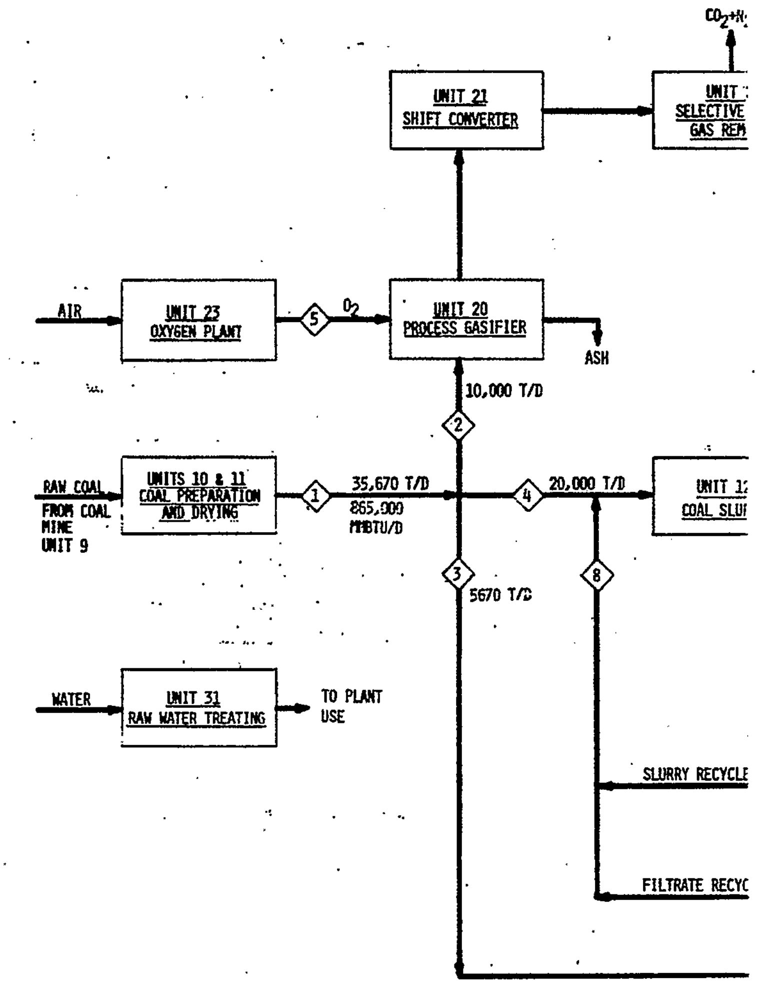


16	17	18	19	20	21	22
BUTANE	LIGHT NAPHTHA	HEAVY NAPHTHA	DIESEL	FUEL OIL	SNG	ALCOHOL MIX
-	-	-	-	-	4	-
-	-	-	-	-	5	-
-	-	-	-	-	106	-
-	-	-	-	-	329	-
-	-	-	-	-	-	14
-	-	-	-	-	2318	-
-	-	-	-	-	-	-
-	-	-	-	-	77	-
-	-	-	-	-	-	-
-	-	-	-	-	137	-
28	-	-	-	-	-	-
144	3	-	-	-	362	-
1	43	-	-	-	-	-
2	196	-	-	-	1	-
-	354	3	-	-	-	-
-	17	549	30	-	-	-
-	-	41	1028	41	-	-
-	-	-	10	321	-	-
-	-	-	-	-	-	269
-	-	-	-	-	-	-
174	614	593	1068	362	3344	283

STREAM NO.	24	25	26	27	28	29
STREAM NAME	CO_2 VENT	OXYGEN	PROCESS WATER	SOUR GAS	STACK GAS	SULFUR
Flow, T/D						
H_2	-	-	-	-	-	-
CO	-	-	-	-	*1000 ppmv	-
CO_2	7282	-	-	-	-	-
CH_4	-	-	10 PPM	-	-	-
H_2S	-	-	4	3.5	1 ppmv	-
N_2	-	-	-	-	-	-
O_2	-	7	-	-	-	-
H_2O	192	-	6125	3	-	-
NH_3	-	-	2	1.5	-	-
COS	-	-	-	-	50 ppmv	-
SO_2	-	-	-	-	-	-
COD	-	-	4	-	-	-
ORGANICS	-	-	1	-	-	-
CHLORIDES	-	-	1	-	-	-
CARBONATES	-	-	5	-	-	-
SOLIDS	-	-	2	-	-	-
TOTAL T/D	7474	7	6144	8	4000	587

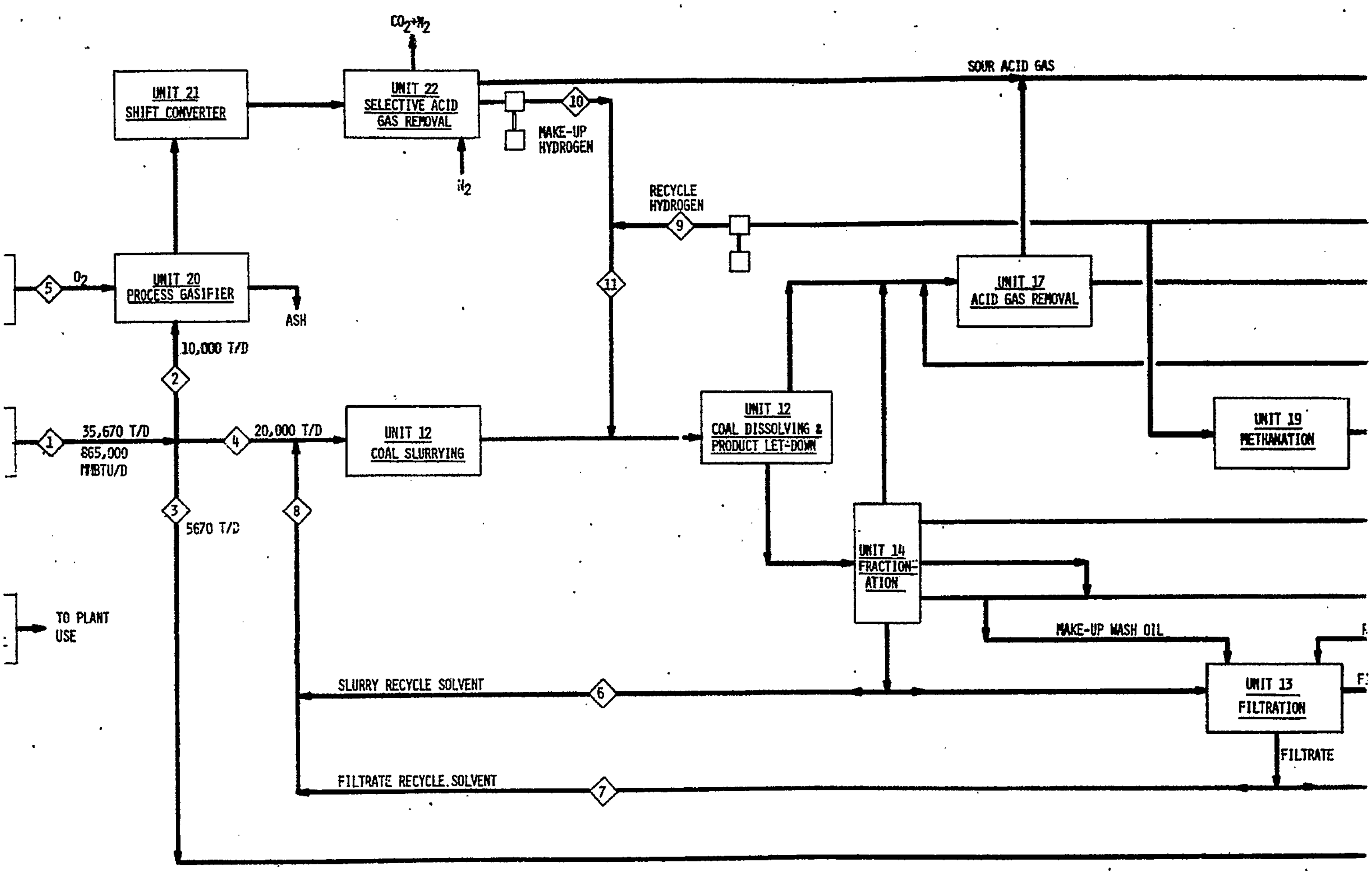
*MAXIMUM

Diagram



CO2+H
UNIT 21
SHIFT CONVERTER
UNIT
SELECTIVE
GAS REM
AIR
UNIT 23
OXYGEN PLANT
5
O2
UNIT 20
PROCESS GASIFIER
ASH
10,000 T/D
2
RAW COAL
FROM COAL
MINE
UNIT 9
UNITS 10 & 11
COAL PREPARATION
AND DRYING
1
35,670 T/D
865,000
MMBTU/D
4
20,000 T/D
UNIT 1
COAL SLU
3
5670 T/D
8
WATER
UNIT 31
RAW WATER TREATING
TO PLANT
USE
SLURRY RECYCLE
FILTRATE RECYC

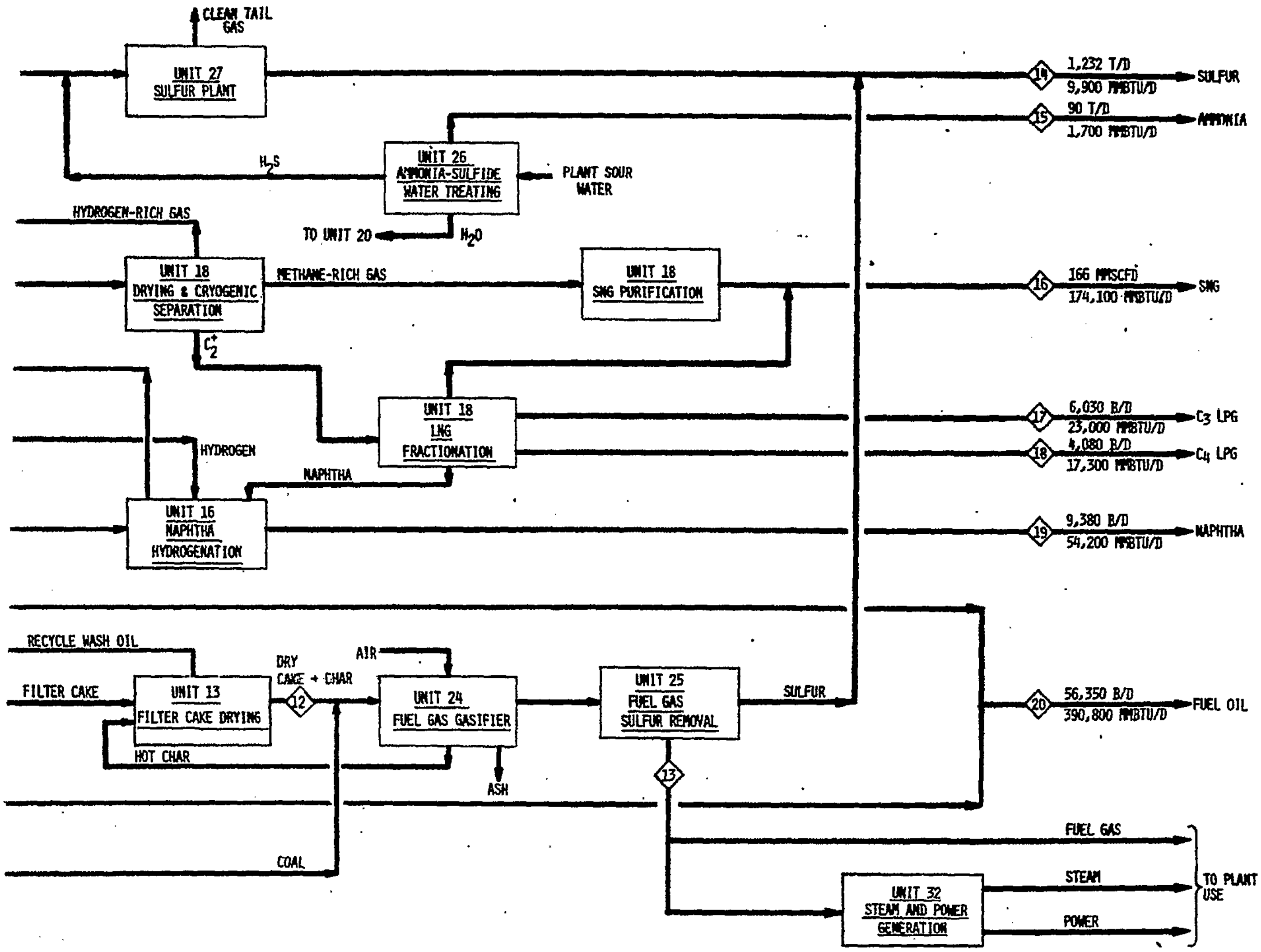
A



STREAM NO.	1	2	3	4	5	6	7	8	9	10
STREAM NAME	CLEAN COAL FEED	COAL TO H.P. GASIFIER	COAL TO L.P. GASIFIER	COAL TO DISSOLVING	OXYGEN TO H.P. GASIFIER	SLURRY SOLVENT	FILTRATE SOLVENT	TOTAL SOLVENT	RECYCLE HYDROGEN	MAKE-HYDRO
O_2, T/D	-	-	-	-	4,418.85	-	-	-	-	-
H_2	-	-	-	-	-	-	-	-	760.50	1,069
N_2	-	-	-	-	78.38	-	-	-	105.91	204
CO	-	-	-	-	-	-	-	-	271.32	625
CO_2	-	-	-	-	-	-	-	-	-	29
NH_3	-	-	-	-	-	-	-	-	-	-
H_2S	-	-	-	-	-	-	-	-	-	11
H_2O	963.09	270.00	153.09	540.00	-	-	-	-	-	9
C_1	-	-	-	-	-	-	-	-	49.27	1,688
C_2	-	-	-	-	-	-	-	-	-	-
C_3	-	-	-	-	-	-	-	-	-	-
nC_4	-	-	-	-	-	-	-	-	-	-
IBP-200	-	-	-	-	-	-	-	-	-	-
200-400	-	-	-	-	-	-	-	-	-	-
400-450	-	-	-	-	-	-	-	-	-	-
450-500	-	-	-	-	-	-	-	-	-	-
500-650	-	-	-	-	-	152.18	84.89	237.07	-	-
650+	-	-	-	-	-	35,701.82	19,915.11	55,616.93	-	-
Residue	-	-	-	-	-	1,485.60	-	1,485.60	-	-
Ash	4,209.07	1,180.00	669.07	2,360.00	-	2,660.40	-	2,660.40	-	-
Coal (MAF)	30,497.96	8,550.00	4,847.96	17,100.00	-	-	-	-	-	-
Sulfur	-	-	-	-	-	-	-	-	-	-
Total, T/D	35,670.12	10,000.00	5,670.12	20,000.00	4,497.23	40,000.00	20,000.00	60,000.00	1,187.00	3,638
MMSCFD	-	-	-	-	107.1	-	-	-	299.3	50
BPD	-	-	-	-	-	-	-	-	-	-
Heating Value, Billion BTU/D	865.0	242.5	137.5	485.0	-	-	-	-	-	-

Figure B-2 Oil/Gas Plant Block Flow Diagram

B



10 MAKE-UP HYDROGEN	11 TOTAL HYDROGEN	12 DRY CAKE (ONLY)	13 PLANT FUEL GAS	14 SULFUR	15 AMMONIA	16 SNG	17 C_3 LPG	18 C_4 LPG	19 NAPHTHA	20 FUEL OIL
-	-	-	-	-	-	-	-	-	-	-
,069.20	1,829.70	-	331.15	-	-	6.80	-	-	-	-
204.70	310.61	-	14,684.08	-	-	204.17	-	-	-	-
625.42	896.74	-	8,533.99	-	-	5.71	-	-	-	-
29.23	29.23	-	1,067.58	-	-	43.83	-	-	-	-
-	-	-	-	-	90.10	-	-	-	-	-
11.71	11.71	-	-	-	-	-	-	-	-	-
9.17	9.17	-	488.48	-	-	-	-	-	-	-
,688.71	1,737.98	-	-	-	-	2,951.04	-	-	-	-
-	-	-	-	-	-	565.20	11.05	-	-	-
-	-	-	-	-	-	162.56	514.44	77.82	-	-
-	-	-	-	-	-	-	6.28	322.02	53.83	-
-	-	-	-	-	-	-	-	5.88	101.11	
-	-	-	-	-	-	-	-	-	1,123.49	294.98
-	-	-	-	-	-	-	-	-	-	532.58
-	-	2.89	-	-	-	-	-	-	-	510.11
-	-	51.59	-	-	-	-	-	-	-	1,511.36
-	-	119.09	-	-	-	-	-	-	-	8,459.93
-	-	1,318.53	-	-	-	-	-	-	-	-
-	-	2,465.51	-	-	-	-	-	-	-	-
-	-	-	-	-	-	-	-	-	-	-
-	-	-	-	1,231.70	-	-	-	-	-	-
3,638.14	4,825.14	3,957.61	25,105.28	1,231.70	90.10	3,939.31	531.77	405.72	1,278.43	11,308.96
506.8	806.1	-	792.92	-	-	165.9	-	-	-	-
-	-	-	-	-	-	-	6,030	4,082	9,379	56,348
-	-	-	114.9	9.9	1.7	174.1	23.0	17.3	54.2	390.8

TOTAL PRODUCTS
671,000 MMBTU/D
77.6% EFFICIENCY

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National Technical Information Service
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