COLLECTED WORK NO. 11

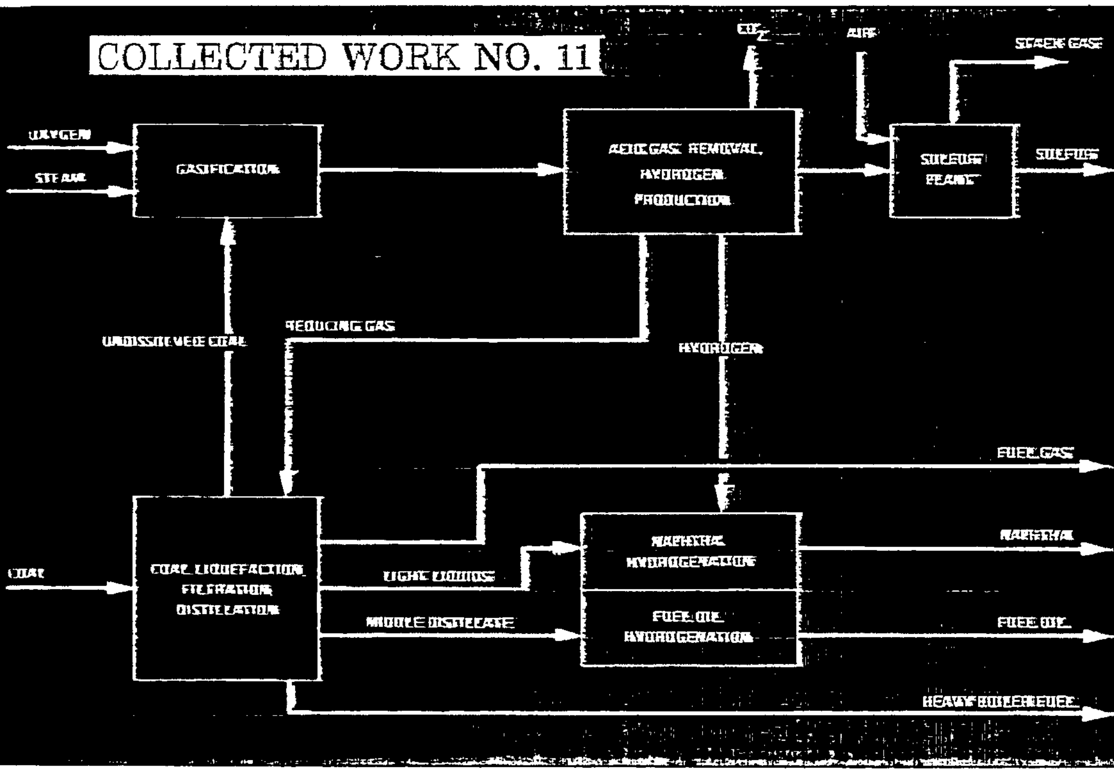


Figure 1. Demonstration plant block flow diagram.

A CEP CAPSULE

Coal Processing:

Producing Clean Boiler Fuel from Coal

Preliminary design for a 10,000 ton/day plant for demonstrating a process that includes modified solvent refining and coal gasification.

J. B. O'Hara, N. E. Jentz, S. N. Rippee, and E. A. Mills
Ralph M. Parsons Co., Los Angeles, Calif.

A study for the Office of Coal Research (OCR) of preliminary design and estimated costs for a demonstration-scale plant to produce clean boiler fuels from coal has been completed by the Ralph M. Parsons Co. Objectives of the preliminary design were as follows:

1. Establish a plant design to effectively produce clean boiler fuels from coal.
2. Estimate fixed capital investment required for design, engineering, procurement, and construction of a coal conversion plant.
3. Estimate the earliest date of production for the coal conversion plant.
4. Estimate budget for each semi-annual period during plant construction.

Process design bases and yields for the plant were provided by OCR process development contractors. The OCR concept with the greatest potential for entirely converting a typical coal into desulfurized liquid fuel was used.

The first step in a program to carry coal conversion into commercial reality, the preliminary design represents extensive engineering research, both for selection of equipment and for processing techniques. The approach accepts potential risks in performance and some cost uncertainty in order to speed development of a viable commercial design. This preliminary design, based on immature technology, precedes experimental results from pilot-plant operations using the two primary coal conversions steps of liquefaction and synthesis gas ("syngas") production.

Demonstration plant objectives were:

1. To speed the arrival of commercial coal conversion processes for production of clean fuels from indigenous high-sulfur coals.
2. To leapfrog the pilot-plant program to gain time in development of commercial processes.
3. To provide enough liquid fuels for prolonged testing in commercial power plants.
4. To define performance requirements and financial incentives for rapid development of large-scale coal conversion plant hardware.
5. To demonstrate operability of commercial-scale plant equipment.
6. To provide a basis for predicting the economics of commercial plants.
7. Following demonstration plant operations, to permit simultaneous design of many commercial plants.

A 10,000 ton/day design

The plant is designed to process 10,000 ton/day of coal. The plant contains two primary process units, a modified solvent refined coal (SRC) unit and a gasifier unit to produce syngas. The gasifier is capable of supplying reducing gases for an SRC-dissolver, an SRC-liquefied coal hydrogenator, and a light oil hydrogenator.

Feed to the gasifier units consists of equal weights of dry filter cake from the SRC unit and filtrate. Phenolics/cresylics produced in the process are recycled to extinction. All effluent streams are treated to meet applicable environmental standards. Enough feed coal is maintained for three days. Enough product is maintained for 30 days.

The process description of the plant contains three major divisions: coal preparation, a coal liquefaction section, and a gasification section. The plant is designed to charge 10,000 ton/day of coal and to produce two low-sulfur liquid fuel streams as major products.

The first product fuel stream is a liquid containing approximately 0.5 wt.-% sulfur, sufficient to fuel a 400-Mw power plant. The second fuel stream is a desulfurized distillate fuel oil product containing 0.2 wt.-% sulfur, sufficient to fuel a 200-Mw power plant. By-products are hydrotreated naphtha and sulfur recovered from the desulfurizing processes. Light hydrocarbons produced are used as plant fuel.

With an alternate product objective, the plant could be converted to produce approximately 65 million cu. ft./day of substitute natural gas (SNG).

The plant receives 12,500 ton/day of coal. Each rail car is dumped into a hopper which can also receive coal from mine trucks. Coal is stockpiled and reclaimed by a bucket wheel, with 900 ton/hr. of stockpiling capacity and 800 ton/hr. for the reclaim system.

Coal is dried and ground before 10,000 tons of coal is mixed with 20,000 tons of recycle solvent. This feed, containing 10,000 ton/day of 1/8-inch-minus coal as a 50-wt.-% slurry, is charged to a reactor where it contacts a reducing gas at about 850°F and 1,000 lb./sq. in. gauge. Following this liquefaction, undissolved coal enters the gasifier plant. Thermal efficiencies are generally high, except for the dissolver at 89.2% and hydrogen purification units at 61%. Estimated overall plant thermal efficiency is 63.5%.

Major equipment costs = $76.8 million

Total major equipment cost for all unit areas is estimated at $76.8 million. Factored total construction costs are estimated to be $195 million. Estimated total fixed capital investment required is $270 million, including total construction costs, home office engineering, escalation and sales taxes. Estimated project total capital required including fixed capital investment, startup costs and recommended working capital amounts to about $310 million.

Estimates are considered within −5% to +20% accurate. They are based on a project duration of 48 months, with January 1, 1974, chosen as engineering start date, construction start about April 1, 1975, and completion and plant startup estimated at late in 1977.

Preliminary conclusions are limited by a comparatively small data base. While considerable data on coal liquefaction exists, conditions of unfiltered dissolver product recycle to form feed coal slurry are based on scant data. Additional work is needed to assure design yields and process operability. Critical design parameters were:

1. Recycle of unfiltered liquid effluent from dissolvers.
2. Hydrogen consumption for the dissolving section as 3 wt.-% of coal feed.
3. Liquid residence time in the preheater and dissolver as one hour.
4. Use of syngas to supply dissolving section hydrogen needs.
5. Solid-to-liquid conversion of coal in the dissolver at 91%.
6. Use of filtration on net dissolver product to remove undissolved solids from product. Filter cake to contain equal weights of undissolved solid and liquid product.
7. Preheater outlet and dissolver temperatures at 900°F and 840°F respectively.
8. Solvent recycle rate at twice the weight of coal feed.

Limited laboratory results indicate use of unfiltered solvent is attractive for both yield and character of liquid coal product, so the demonstration plant is designed using it. Additional data on residence time required to achieve coal liquefaction is needed, but it is logical that residence time could be reduced with higher temperatures and possible pressures.

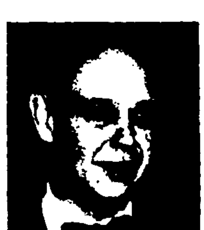
The design work resulted in definition of recommended additional data that should be developed to further assure satisfactory plant performance.

Design analysis showed the possibility of producing desulfurized liquid fuels using an alternative approach to syngas production. The alternative is using liquefaction plant offgases for syngas production rather than plant fuel. The present design employs coal residue gasification from the liquefaction plant for this purpose.

Detailed demonstration plant design could be under way while a pilot plant is operating. Many questions should be resolved by pilot-plant performance along with adjustment of program schedules to include pilot-plant data in demonstration plant design. Figure 1 provides a block flow diagram of a demonstration plant. #

The complete manuscript of 29 pages may be obtained for $6.00 (U.S. prepaid only) by writing AIChE Publications Dept., 345 E. 47th St., New York, N.Y. 10017. Foreign add $2.50, prepaid only. Request MS No. 5745.

O'Hara



Jentz

Rippee

Mills

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COLLECTED WORK NO. 12

COMMERCIAL COMPLEX CONCEPTUAL DESIGN/ECONOMIC ANALYSIS

OIL AND POWER BY COED BASED COAL CONVERSION

R & D REPORT NO. 114 - INTERIM REPORT NO. 1

Contract No. E(49-18)-1775

Date Published: September 1975

Prepared for

ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION

WASHINGTON, D. C. 20545

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SECTION 1

INTRODUCTION

This report presents the results of a preliminary design and economic evaluation for a commercial complex to mine high-sulfur coal and produce low-sulfur synthetic crude oil (syncrude), electrical energy, and sulfur using COED-based pyrolysis technology for the coal conversion portion of the complex.

This work was performed for the Energy Research and Development Administration (ERDA) - Fossil Energy, whose support and guidance in these activities are gratefully acknowledged. From the experience gained during pilot plant operation, the design basis was developed in cooperation with representatives of ERDA and the process development contractor, FMC Corporation.

1.1 OBJECTIVES

The objectives of the work described in this report were to:

(1) Review the experience obtained during the successful operation of the ERDA-supported COED pilot plant operated by FMC Corporation at Princeton, New Jersey, over the period 1970 through 1974.

(2) Develop a conceptual design for a commercial COED-based industrial complex including all operations required to mine coal, prepare it by cleaning and washing it, convert it to ecologically clean liquid and gaseous fuels, and convert the gaseous fuels to electrical energy for sale.

(3) Estimate the economics for the facility to serve as a guide in making decisions regarding future commercial applications of this technology.

(4) Provide recommendations regarding additional development effort to foster commercial exploitation of the technology.

(5) Define probable project and financial parameters for design, engineering, procurement, construction, and startup of the complex.

1.2 REPORT ORGANIZATION

A summary of the material contained in this report is presented in Section 2 to aid the reader in rapid assimilation of its contents.

Sections 3 and 4 provide an introduction and orientation for the detailed design information which follows in later sections. The design parameters and design bases used are summarized in Section 3. Section 4 describes the scope and major units included in the complex. An overview of the method of assembling the principal process units and material flows, presented in the form of a block flow diagram, is also included in Section 4, together with an artist's rendition of the complex. Section 5 contains detailed descriptions of the design, and process flow diagrams are presented in Section 6. The overall materials and energy efficiencies are summarized in Sections 7 through 9.

Section 10 summarizes the major equipment items required; this provides a basis for the fixed capital investment and operating cost estimates which follow. Environmental factors that must be considered are detailed in Section 11. The estimated economics for this type of complex is developed in Section 12. Here the fixed capital investment, other capital requirements, operating requirements and operating cost, and projected profitability are presented. Sensitivity factors are also given.

Opinions regarding projected performance of this facility are presented in Section 13 and, finally, recommendations for future design improvements are given in Section 14. These latter two sections summarize the results of an after-the-fact evaluation of probable performance and suggestions for further design work.

SECTION 2

SUMMARY

A conceptual design and economic evaluation for a project to design, engineer, procure, construct, and start up an industrial complex which will mine high-sulfur coal and convert it to low-sulfur syncrude plus electrical energy has been completed. The results are summarized in this report.

This work was done with the support and guidance of the Energy Research and Development Administration - Fossil Energy. The design basis was developed in cooperation with representatives of ERDA and FMC Corporation, the process developer. The design basis utilized data and experience gained during the pilot plant operation.

The scope of the industrial complex consists of a large captive coal mine supplying the feed material to a coal preparation plant, which in turn supplies approximately 25,000 TPD of clean, washed coal to a COED-based pyrolysis coal conversion plant. In the COED facility, the feed coal is converted to 25°API, 0.1% sulfur syncrude plus low-sulfur fuel gases, as well as byproduct sulfur. The fuel gases are fed to a close-coupled electrical power generation plant which produces electricity for export; it also produces steam for captive use in the complex. The complex is a grass roots facility and is conceived to be located in the Eastern Region of the Interior Coal Province. The mine-mouth processing facility meets desired location criteria consisting of significant resource of high-sulfur coal with a large utility/industrial market nearby, with ecological restrictions for direct consumption of the indigenous high-sulfur coal.

This design provides the equipment and operating flexibility to process feed coal with a range of analyses which might be expected over the course of a 20-year operating life, using coal typically mined in the Eastern Region of the U.S. Interior Coal Province. This distinguishes the design from other designs which have been based on a single typical coal analysis and which might be called single feed source or "point" designs. The use of a fixed coal feed rate and variable coal characteristics requires higher fixed capital investment to provide the necessary flexibility; it also results in variable product rates.

The process flowsheets and accompanying heat and material balances which are presented are based on a typical coal analysis which is intermediate between the extreme analyses that might be encountered during the project life. All equipment was sized to handle the range of feed analyses and resulting operating condition adjustments to permit operation at capacity approximately 95% of the onstream time.

For the typical design case, approximately 36,000 TPD of run-of-mine (ROM) coal is mined and processed through a coal preparation plant to produce approximately 27,500 TPD of clean, washed coal feed to the COED pyrolysis plant. Products from the process plant include 28,000 bbl/day of 25°API syncrude with a maximum sulfur content of 0.1% by weight plus about 830 MW of electrical power for sale. About 750 long tons of sulfur is produced as by-product.

Considerable attention was given to methods of scale-up; the scale-up factor from pilot plant to the conceptual design reported here was of the order of 700. Methods of scale-up were selected to provide efficiency, operability, and process controls.

To develop the conceptual design it was necessary to use extrapolation procedures in certain cases. For these cases, the basic chemical engineering phenomena were examined in detail with the objective of developing a sound basis for the extrapolation and scale-up. An example is that the conceptual design encompasses gasification of pyrolysis-produced char to consume a total of approximately 98% of the carbon contained in the feed coal. This represents an example of extrapolation; in a typical pilot plant run approximately 30% of the carbon was converted in the pyrolysis section. About 6% of the carbon in the resulting pyrolysis-produced char was gasified by reaction with steam and oxygen. This compares with 66% steam-oxygen gasification required in the conceptual plant design described here.

The design represents an assessment of a proposed configuration and potential economics for this type of technology. To accomplish this objective required the use of engineering judgement for the scale-up and the selection of the equipment required to achieve the stated objectives. It also represents an exposition of factors required to integrate the coal conversion plant with a closely coupled electrical power generation facility and a large coal mine. The development of the interfaces between the coal mine, process plant, and power plant has defined a number of the design and operational options which exist for maximizing efficiencies and profitabilities.

Approximately 500 acres should be allocated for the complex site exclusive of the coal mine. Over a 20-year project life, about 42 square miles would be mined. An artist's conceptual drawing of the complex is presented.

The estimated fixed capital investment for the complex is $1 billion; all estimates are in first-quarter 1974 dollars. The total capital investment is estimated to be $1.125 billion; this includes the cost of initial raw materials, catalysts and chemicals, allowance for startup and land acquisition and initial working capital.

The population of the complex is estimated to be about 1700. Operating costs are projected to be about $127 million per year. The required plant revenue for a 10% discounted cash flow rate of return (DCF) with 65% debt at 9% interest is $300 million per year. Typical required selling prices for the mixed product slate at 10% DCF, after by-product sulfur credit, are as follows:

Syncrude, $/bbl	Electricity, mils/kW-hr
10	32
15	25
18	20
26	10

Other cases and sensitivities of required selling prices and profitability to key economic parameters are presented.

A representative project schedule for the design, engineering, construction, and startup is given; a 57-month schedule to mechanical completion is projected, and a probable fund drawdown schedule is presented.

The conceptual commercial COED process plant described herein has been designed to be capable of processing the design feed at the design rate and produce products of design quality and quantity. Where uncertainty in basic information existed, the equipment has been specified to cover this uncertainty.

The design is considered to be workable with the understanding that the estimated costs that are reported here have the probability of being greater than if additional information were available; this is often the case for first-generation plants.

The design of the char gasifiers for high carbon conversion represents an extension of the COED pilot plant experience. Prior experience obtained in the first-stage gasification step in the pilot plant, along with additional available kinetic data, were used to design the commercial-scale units. The results were compared with small-scale experimental work using COED char and a resulting kinetic model developed by the Institute of Gas Technology (IGT) to correlate the data. In addition, IGT was authorized to conduct a two-phase experimental program to investigate conditions required to achieve the specified gasification results. This program was performed under Parsons Subcontract No. 4-SC-5054-3 and consisted of:

Phase I: Thermogravimetric Study.

Phase II: Fluidized-Bed Reactor Gasification in a 6-Inch Diameter Reactor.

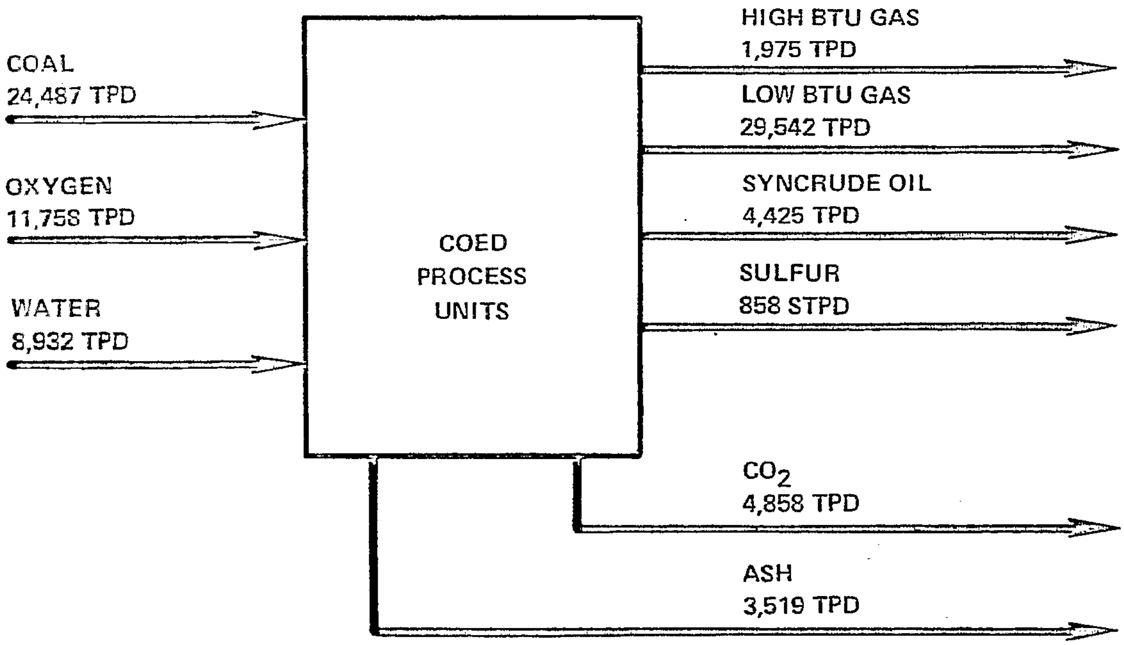
The results of the IGT study indicated:

(1) Using the gasifier reactor sizes specified, the bed temperature should be increased to 1,820°F.

(2) Additional experimentation should be conducted with the 6-inch gasifier at temperatures to 1,800°F; particular attention should be given to the determination of fluidization velocities necessary to inhibit sintering at the elevated temperature.

Available experience indicates that the beds can be successfully operated at the 1,800°F level. The gasifiers have been designed to permit operation at that temperature.

A number of potential design improvements are presented. These improvements would be expected to improve the economics when successfully reduced to practice.



TOTAL IN = OUT = 45,177 TPD

Figure 7-1 - Overall Material Balance -
Typical Operation for Process Sections

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COLLECTED WORK NO. 13

PRELIMINARY ECONOMIC ANALYSIS OIL AND POWER BY COED-BASED COAL CONVERSION

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CHAPTER XII

--PRELIMINARY ECONOMIC ANALYSIS--
OIL AND POWER BY COED-BASED COAL CONVERSION

J. B. O'Hara
R. V. Teeple

I. INTRODUCTION

The development of viable processes to convert high sulfur coal to environmentally clean liquid fuels has become a U.S. objective, with the Energy Research and Development Administration-Fossil Energy (ERDA-FE) primarily responsible for achieving this goal. A number of processes are under study. General classifications include:

- Pyrolysis, in which the coal is heated to a high temperature to produce a gas, an oil, and a char.
- Hydroliquefaction, in which coal in the form of a slurry is contacted with hydrogen at an elevated pressure and temperature to produce a liquefiable product with reduced sulfur content.
- Donor solvent extraction, in which a hydrogenated coal-derived liquid is contacted with coal at an elevated temperature and pressure to extract a major proportion of the coal substance and produce an oil.
- Indirect liquefaction, in which coal is gasified to produce a mixture of carbon monoxide and hydrogen (syngas), contaminants such as hydrogen sulfide and solid particulates are removed, and liquids are produced by catalytic reaction of the syngas at an elevated temperature and pressure.

For each of the process categories listed, variations are possible. Also, in all cases some substitute natural gas (SNG) may be coproduced.

A key goal of the development program is to define capabilities and projected economics of the separate technologies. As one step in this program, The Ralph M. Parsons Company has prepared a conceptual design and an economic evaluation for a commercial complex to mine approximately 36,000 tons per day (TPD) of high-sulfur coal and produce low-sulfur synthetic crude oil (syncrude), electrical energy, and sulfur using COED-based pyrolysis technology for the coal conversion portion of the complex. COED is a multistep low-pressure coal pyrolysis process; the pyrolysis portion of the technology plus gasification of a part of the char produced in the pyrolysis steps were successfully tested in pilot plant operations at Princeton, New Jersey from 1970 to 1974. The development program was sponsored by the Office of Coal Research, now a part of ERDA-FE. FMC Corporation was the development contractor.

This paper briefly describes key elements of the conceptual design and then summarizes the projected economics. Economic standards used in this presentation conform to the uniform practice procedure established by the symposium sponsors. These economic standards are shown in the Appendix to this paper.

II. PARSONS ROLE

The COED-based conceptual design/economic analysis is one of a number of task assignments to be completed by Parsons under contract to ERDA-FE.

Parsons is actively assisting ERDA to develop viable commercial plants for the conversion of coal to environmentally clean fuels. There are two distinct elements involved in this work:

(1) Preliminary design services, in which Parsons develops preliminary/conceptual designs and economic evaluations for large-scale coal conversion plants. Examples of completed tasks are a demonstration plant design to produce approximately 25,000 barrels per day of clean

boiler fuels, which was published as OCR R&D Report No. 82, Interim Report No. 1,[1] and the commercial complex to produce oil and power by COED-based coal conversion discussed in this paper. The full report appeared as ERDA R&D Report No. 114, Interim Report No. 1.[2]

Under this contract, Parsons will also develop conceptual designs for a Fischer-Tropsch plant to be responsive to U.S. requirements, an oil/gas plant, a coal-oil-gas (COG) multiproduct facility, a commercial solvent refined coal (SRC) facility, and a multiunit demonstration facility. Each of these designs will include captive coal mines. A parametric economic analysis will be prepared for each design.

(2) Parsons also supplies technical evaluation contractor services to assist ERDA in monitoring certain of the liquefaction development programs.

III. OBJECTIVE

The primary objective of this paper is to summarize the projected economics for a commercial-scale coal conversion complex using COED-based pyrolysis technology to produce syncrude plus electrical power as the principal products.

IV. FACILITY: DESIGN BASIS

A Design Basis describing design philosophy, key data to be used for design, and principal process steps in the pyrolysis-gasification section was cooperatively developed by the client (OCR), the process developer (FMC Corporation), and the designer (Parsons) prior to undertaking the detailed conceptual design effort. Key design parameters for the complex are shown in Table 1.

A most important characteristic of the design is that it provides the equipment and operating flexibility to process feed coal with a range of analyses that might be expected over the course of a 20-year operating life, using coal typically mined in the Eastern Region of the U.S. Interior Coal Province.

Table 1 - Key Design Parameters

Parameters	Design Basis
Pyrolysis Plant Feed Rate	21,500 TPD moisture and ashfree coal
Primary Products	Syncrude with 0.1%S, maximum electrical energy
Secondary Product	Sulfur: 99+% pure
Location, Site Conditions	Eastern Region, Interior Coal Province
Scope	Grass-Roots Coal Conversion and Power Generation Facility with an adjacent captive coal mine
Feed Coal Composition	Range of analyses expected over a 20-year operating life

This distinguishes the design from others that have been based on a single typical coal analysis and that might be called single-feed source or point designs. The use of variable feedstock characteristics requires higher fixed capital investment to provide the necessary flexibility. It also results in variable quantities of products. These factors will be referred to during the course of subsequent discussions of the characteristics and economics of the process.

In order to describe the design, process flowsheets and accompanying heat and material balances have been based on a typical coal analysis that is intermediate between the extreme analyses that might be encountered during the project life. The equipment was sized and operating ranges were provided to permit capacity operation with any of the feed coal analyses, 95% of on-stream time.

The design and economics to be described here represent one potential application of COED technology. COED pyrolysis produces a gas, a liquid, and a significant amount of char. There are a number of alternatives possible, depending upon what is done with the char. In addition to the case where it is converted to electrical energy, the char can be used as raw material for SNG manufacture, for methanol or ammonia manufacture, or for liquid fuels production using Fischer-Tropsch technology.

V. FACILITY DESCRIPTION

A simplified block flow diagram for the complex is shown in Figure 1. Principal elements of the complex consist of units to accomplish the following objectives:

(1) Mine coal to supply approximately 36,000 tons per day of run-of-mine (ROM) coal.

(2) Crush and wash coal to minus 1/8-in. size.

(3) Dry coal.

(4) Pyrolyze coal in a series of fixed fluidized beds using heat generated when gasifying the residual char.

(5) Condense oil from the pyrolysis vapors.

(6) Remove fine solids from the pyrolysis oil; filtration is used in this design.

(7) Hydrotreat recovered oil to reduce viscosity as well as sulfur, nitrogen, and oxygen contents.

(8) Desulfurize pyrolysis gas to produce a clean fuel and recover elemental sulfur.

(9) Gasify pyrolysis-produced char to generate low-Btu fuel gas, and desulfurize it to produce a clean fuel and recover elemental sulfur.

(10) Produce oxygen for use in the gasification step.

(11) Produce hydrogen for use in the hydrotreating step.

(12) Produce electric power from the pyrolysis and low-Btu fuel gases.

(13) Treat plant waste streams.

(14) Provide offsite and ancillary facilities for efficient operation of the complex.

An artist's conceptual drawing of the complex is shown in Figure 2. In it is seen the coal feed to the coal preparation plant at the upper right-hand portion

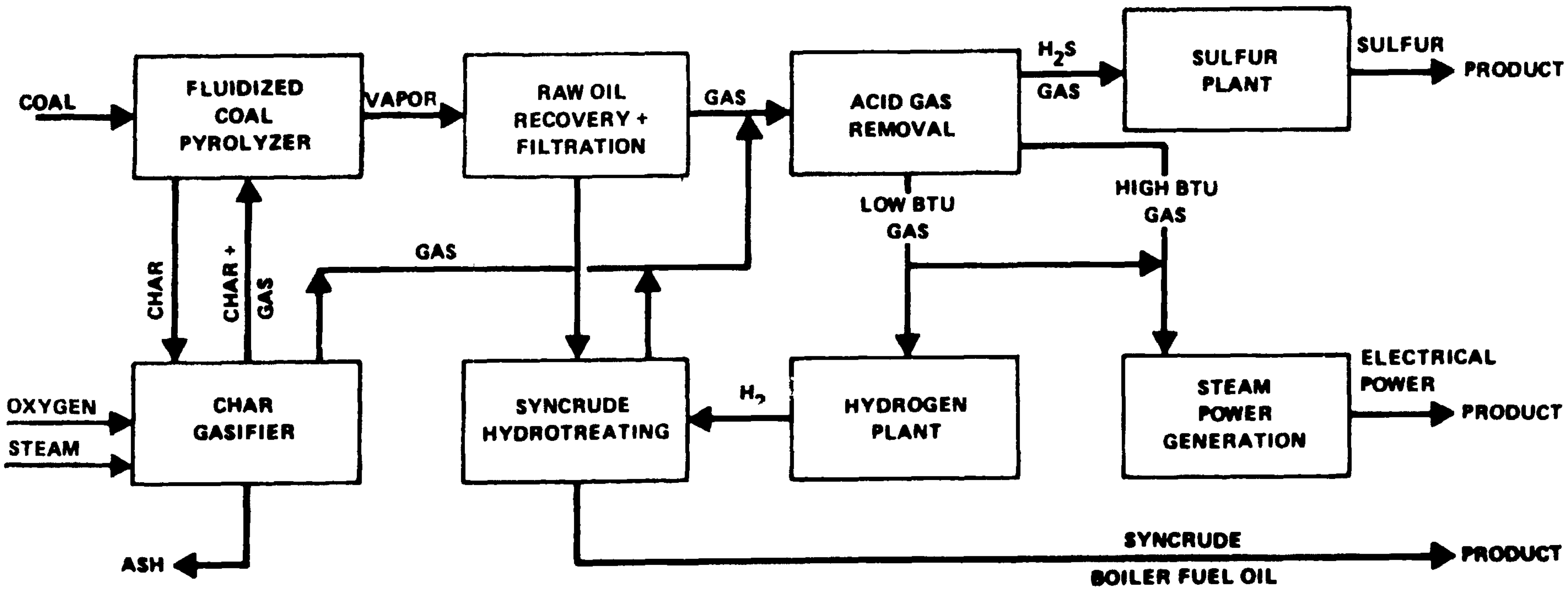


Figure 1 - Simplified Block Flow Diagram

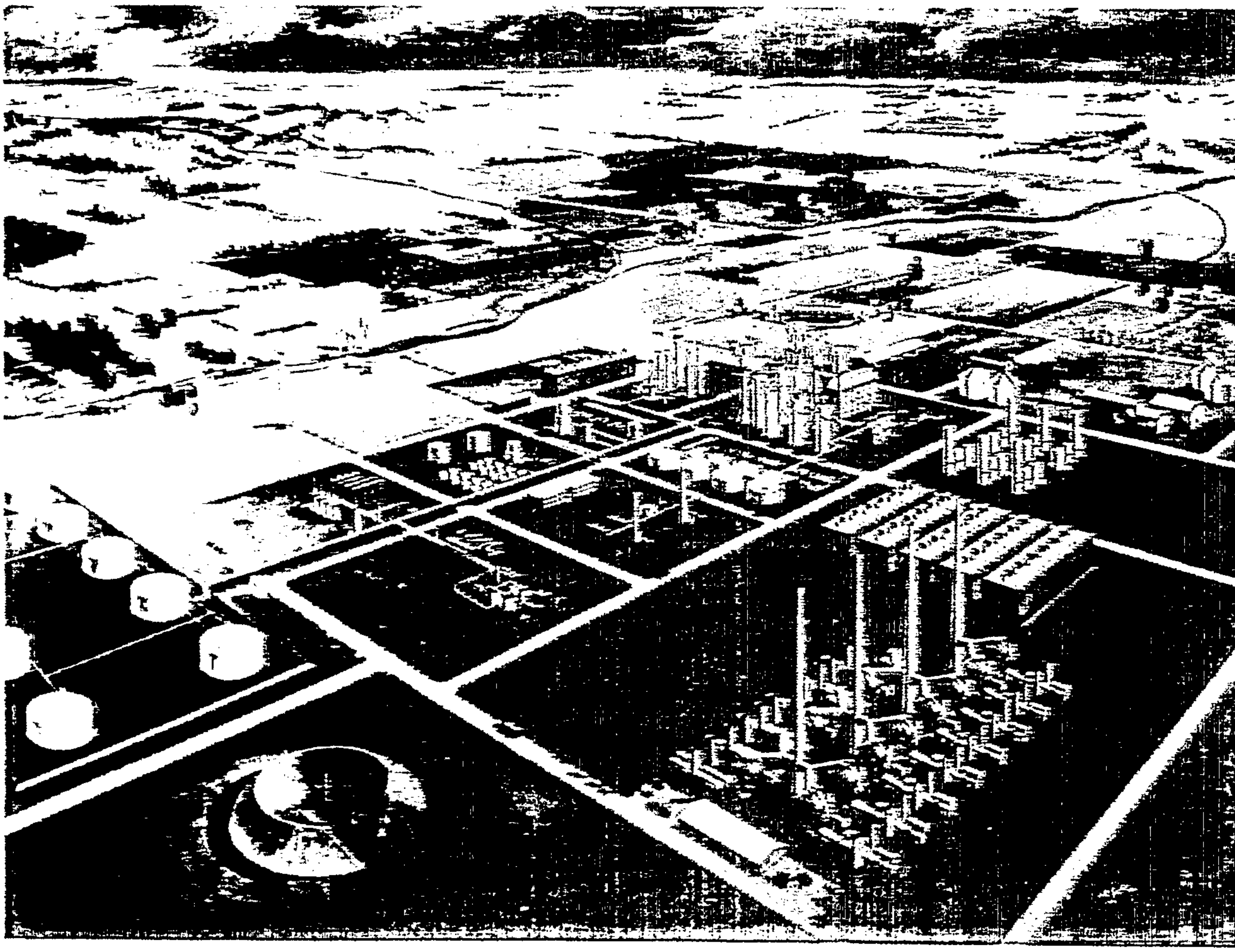


Figure 2 - Artist's Concept

of the drawing. Other key units include the major process areas of pyrolysis and char gasification at the upper left-hand side, the oil hydrotreating section that produces the syncrude at the lower left-hand side, and the power generation plant at the lower right-hand side. Approximately 500 acres should be allocated for the complex, exclusive of the coal mine.

A brief description of the processing follows.

The process complex is supported by an integrated strip mine. Basic mine plant criteria include an average seam thickness of 5 ft, an average overburden depth of 60 ft, and a deposit sufficient for a 20-year project life. Over the project life, approximately 26,500 acres, equivalent to nearly 42 square miles, would be mined out.

The mining plan consists of operating three separate mine faces in proximity to each other. The production rate of ROM coal will be approximately 36,000 tons per stream day (TPSD).

The coal preparation unit receives ROM coal and conditions it to be suitable for feed to the pyrolysis unit. Conditioning consists of grinding and washing to produce, typically, approximately 27,500 TPSD of cleaned coal of -6 mesh size and including about 5% free moisture, 5.5% inherent moisture, and 10.9% ash.

Coal composition will vary during operation of the complex. Ranges of composition expected, stated as cases A and B, are shown in Table 2.

These were selected after analyses of coal compositions in the Eastern Region of the U.S. Interior Coal Province, and represent the 95% probable limits of volatile matter, moisture, and ash contents to be expected. An intermediate composition, referred to as a typical case, is also shown; this was chosen as the basis for the sample heat and material balances shown in the material and energy balances to follow. Equipment was sized and process conditions chosen so that the complex would operate at design capacity 95% of the on-stream time.

Coal received from the cleaning and crushing plant is dried by contact with hot low-Btu gas produced in the char gasifiers and then fed to the pyrolyzers. Pyrolysis is accomplished in three fluidized beds operating at successively higher temperatures, but all at essentially the same pressure (about 15 psig). Typical operating temperatures for the pyrolyzers are 575°F, 815°F, and 1050°F. Heat for the pyrolyzers is supplied from the char gasification step; a portion from the low-Btu gas, and a portion from hot recycle char.

Char produced in the pyrolyzers is gasified by contact with steam and oxygen. Design conditions are to convert 98% of the carbon in the coal fed to the complex. Gasification is accomplished in three stages.

Table 2 - Feed Coal Compositions

Composition	Bases	Compositions (wt %)		
		Case A	Case B	Typical Case C
Proximate				
Moisture	wet	9.5	11.5	10.5
Ash	dry	10.45	13.95	12.2
Volatile Matter	MAF	44.15	48.65	46.4
Fixed Carbon	MAF	55.85	51.35	53.6
Ultimate				
Carbon	MAF	79.8	77.4	78.6
Hydrogen	MAF	5.2	5.6	5.4
Nitrogen	MAF	1.5	1.5	1.5
Sulfur	MAF	4.3[a]	4.3[a]	4.3[a]
Oxygen	MAF	9.2	11.2	10.2
		100.00	100.00	100.00

[a]Design handles sulfur range of 2.8% to 5.8% for each case.

The ash from the gasifiers is cooled, moistened, and returned to the mine for backfill. The gas produced is rich in hydrogen and carbon monoxide; it is cooled as it dries the feed coal and passes through the pyrolyzers. It is desulfurized and then used as fuel to the power plant. A portion, about one-eighth, is separated from the power plant feed stream and used as feedstock for the manufacture of hydrogen for use in the oil hydrotreater.

Raw COED oil and gaseous products are generated as a vapor stream in the pyrolyzers. Entrained dust is removed by cyclones, the gas-vapor mixture from the three pyrolyzers is combined and fed to an oil recovery tower. The oil is recovered as several fractions of hot liquid. Uncondensed product gas is compressed, treated to remove the sulfur-containing contaminants, and then fed to the power plant as fuel.

Heavy and intermediate fractions of oil are combined and filtered to remove solid particles. The filtered oil is hydrotreated in a process somewhat similar to petroleum hydrotreating. The result is a low-sulfur syncrude.

Fuel gases produced in the pyrolysis and gasification units are combined and sent to the power and steam-generating plant that consists of 13 large gas turbine-generator-boiler packages. These units produce steam and electric power needed to operate the complex plus electric power for sale. For the typical design case, approximately 5 million pounds per hour of 600-psia steam are required.

Product rates from the complex are shown in Table 3. Included are the typical product slate using typical coal composition feed as well as maximum and minimum oil production cases based on coal composition differences shown in Table 2.

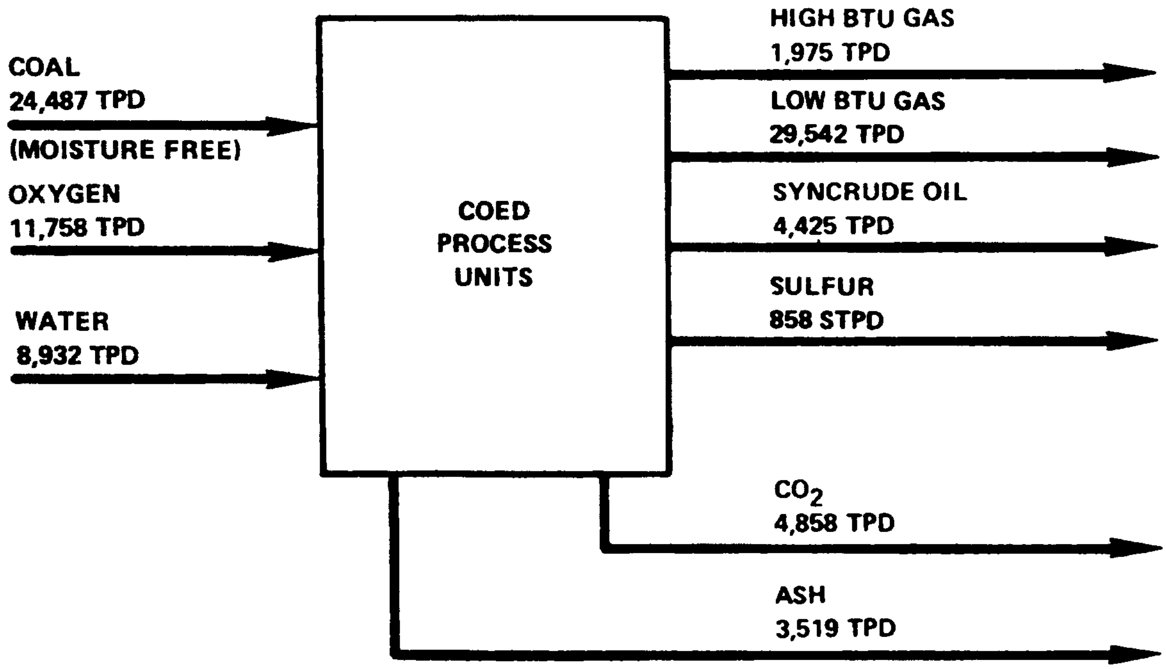
The plant is designed to meet environmental standards. Procedures used to treat gaseous effluents with predicted quantities and compositions of the

Table 3 - Product Production Rates

Products	Characteristics	Production Rates		
		Maximum Oil	Minimum Oil	Typical Oil
Syncrude	Approx 25°API S < 0.1% wt	32,750 BPD	24,000 BPD	28,000 BPD
Electrical Power	138 kV	270 MW	1,150 MW	830 MW
Sulfur	99.9% pure	825 STPD	880 STPD	858 STPD

effluents have been published.[3] Facilities for treatment of liquid and solid discharges are described in the full ERDA report.[2]

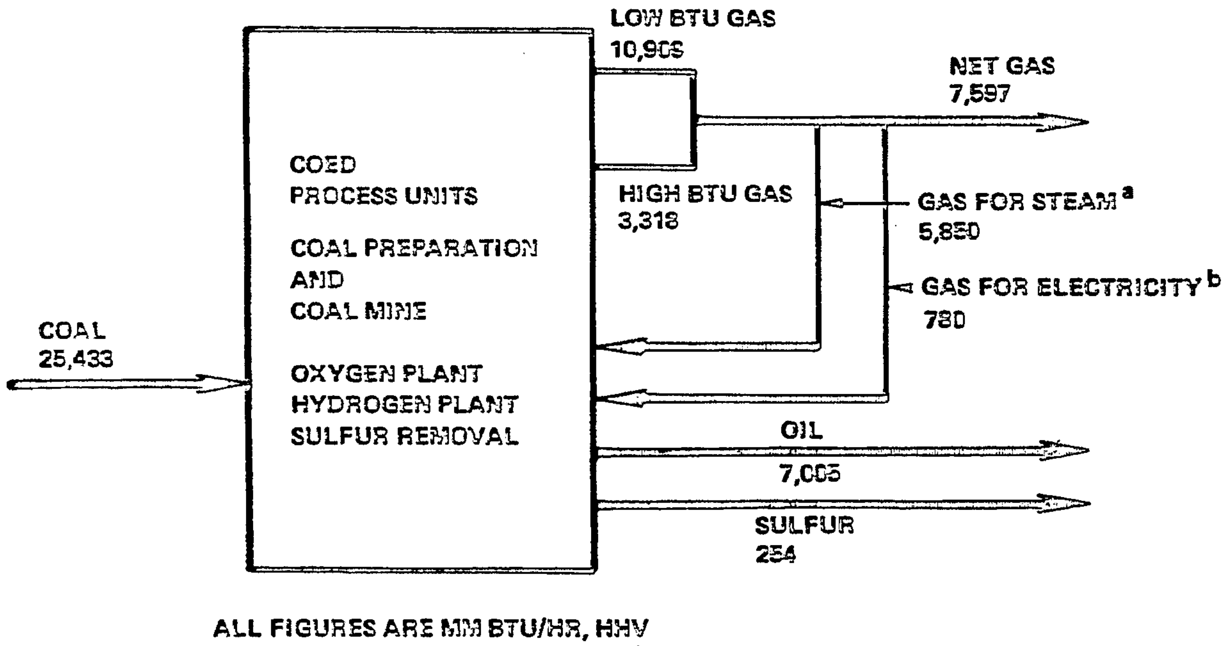
The overall material balance for the process sections is shown in Figure 3 for the typical case. Here it is seen that approximately 24,500 TPSD of moisture-free coal produces about 4,500 TPSD of syncrude plus approximately 2,000 TPSD of high-Btu (890 Btu/CF) gas and about 30,000 TPSD of low-Btu (250 Btu/CF) gas, also about 850 TPSD of sulfur.



TOTAL IN = OUT = 45,177 TPD

Figure 3 - Overall Material Balance

Predicted thermal efficiency of the process and supporting units is depicted in Figure 4. Thermal efficiency of the operation to convert coal to syncrude, feed gas for the power-steam generation plant, and sulfur is estimated to be about 58%. Also of interest is the energy distribution for the total complex,



$$\text{THERMAL EFFICIENCY} = \frac{7{,}597 + 7{,}005 + 254}{25{,}433} = 58.4\%$$

a) GAS FOR IN-HOUSE STEAM REQUIRED: $\dfrac{4{,}970}{0.85_{\text{EFF}}} = 5{,}850$

b) GAS FOR IN-HOUSE ELECTRIC POWER REQUIRED: $\dfrac{80 \text{ MW} \times 3{,}413}{0.35_{\text{EFF}}} = 780$

Figure 4 - Thermal Efficiency Based on Export of Fuel Gas

including electrical power generation shown in Table 4. Here we see that energy in the products represents about 40% of the feed coal energy; conversely, about 60% of the energy in the feed coal is consumed in the manufacturing process. Obviously, a significant amount of energy is dedicated to production of electrical energy from clean fuel gases; 35% was used for the efficiency of the electrical production step from fuel gas.

VI. ECONOMIC ANALYSIS BASIS

The economic basis prescribed for the symposium, and shown in the Appendix to this paper, was used for the development of the economics to be reported in the

Table 4 - Energy Balance for Total Complex

Energy Distribution	Million Btu/hr HHV	Percent
Energy Source: Coal	25,433	100.0
Energy Consumed in Manufacturing		
Mining and Preparation	551	2.2
Oxygen	2,682	10.6
COED	5,707	22.4
Power House	6,389	25.1
Total	15,329	60.3
Energy Value of Product		
COED Oil	7,005	27.5
Export Power	2,823	11.1
Sulfur	276	1.1
Total	10,104	39.7

following sections to provide conformance to a standard comparison base. For reference, the original estimates[2] were in first quarter 1974 dollars and essentially on a fully costed basis involving independent estimates for each of the economic input items.

For this design, the process plant feed coal, power, steam, and oxygen costs are internal costs; all utilities required to operate the facility are produced within the complex.

The complex has been divided into five cost center modules for the purpose of this analysis. They are described in Table 5.

VII. ECONOMIC ANALYSIS

A. Fixed Capital Investment

A summary of the estimated Fixed Capital Investment is shown in Table 6. The total, adjusted to a mid-1975 basis, is approximately 1.3 billion dollars.

B. Total Capital Investment

The Total Capital Investment, by module, is shown in Table 7. In addition to Fixed Capital Investment, estimated costs of initial charges for catalysts and chemicals, startup costs, working capital, and land/rights-of-way are also indicated. Also included is the percentage of capital-cost distribution for each of the modules. The highest cost modules, in decreasing order of magnitude, are the process area, power plant, and coal mine. Total Capital Requirements are approximately 1.4 billion dollars.

C. Operating Costs

A summary of the annual operating costs is shown in Table 8. The total is approximately $185 million per year. Module costs, in decreasing order of magnitude, include highest costs for the coal mine area followed by the process area and the power plant.

Table 5 - Module Descriptions

Module Number	Title	Components
1	Coal Mine	Coal Mine, Primary Crusher, Transport to Coal Preparation Plant
2	Coal Preparation Plant	Coal Preparation Plant, Transport to Process Plant
3	COED Process Facilities	Coal Drying, Pyrolysis, Gasification, Acid Gas Treatment, Oil Recovery, Filtration, Oil Hydrotreating, Hydrogen Plant, Oxygen Plant
4	Power Plant	Fuel Gas Compression, Power Plant
5	Offsites	Support Facilities, Feedwater Treatment, Waste Treatment

Table 6 - Estimated Fixed Capital Investment

Modules	Capital ($ Million)
Coal Mine	122.8
Coal Preparation	34.0
Process	488.0
Power Plant	441.0
Offsites	90.4
Total Constructed Costs	1,176.2
Home Office Cost	75.0
Sales Tax	23.0
Total Fixed Capital Investment	$1,274.2
Say	$1,275.0

Table 7 - Total Capital Costs
($ Million)

Item	Coal Mine	Coal Prep	COED Process	Power Plant	Offsites	Total
Fixed Capital Investment	132.5	37.0	529.0	477.0	99.5	1,275.0
Initial Catalyst and Chemicals	-	-	7.0	-	-	7.0
Startup Cost @ 2% of Facilities	2.7	0.8	10.5	9.5	2.0	25.5
Depreciable Investment	135.2	37.8	546.5	486.5	101.5	1,307.5
Working Capital (60 days)	26.3	2.5	38.1	28.1	7.8	102.8
Land, Rights of Way @ $5000/acre	-	-	-	-	-	2.5
Total Capital Requirements	161.5	40.3	584.6	514.6	109.3	1,412.8
Percent of Total	11.5	2.9	41.4	36.5	7.7	100

Table 8 - Annual Operating Cost Summary
($ Million/Year)

Cost Item	Coal Mines	Coal Preparation	Process	Power Plant	Offsite	Total
Mine Royalty - @ $1.50 per ton	13.6	-	-	-	-	13.6
Materials and Supplies						
Operating Supplies	3.6	0.2	2.6	-	0.5	6.9
Equipment Operation	13.4	-	-	-	-	13.4
Maintenance Materials (2% of facilities)	-	0.7	10.6	9.5	2.0	22.8
Catalysts and Chemicals	-	-	5.9	-	2.7	8.6
Water - @ $0.05 per thousand gallons	-	-	-	-	0.7	0.7
Total Materials and Supplies	17.0	0.9	19.1	9.5	5.9	52.4
Labor						
Operating Labor and Supervision	8.2	0.2	2.9	1.6	0.3	13.2
Maintenance Labor and Supervision	1.6	0.7	10.1	9.2	1.9	23.5
Payroll Burden - @ 20% of labor	2.0	0.2	2.6	2.1	0.4	7.3
Plant Overhead - @ 20% of labor	2.4	0.2	3.1	2.6	0.5	8.8
Union Welfare - @ $1.55 per ton	14.1	-	-	-	-	14.1
Total Labor Costs	28.3	1.3	18.7	15.5	3.1	66.9
G&A Overhead - @ 2% of facilities	2.7	0.7	10.6	9.5	2.0	25.5
Miscellaneous Costs	0.4	-	-	-	-	0.4
Insurance - @ 2% of facilities	2.7	0.7	10.6	9.5	2.0	25.5
Total Operating Costs	64.7	3.6	59.0	44.0	13.0	184.3
Percent of Total	35	2	32	24	7	100

To develop the estimates just presented, the first quarter 1974 estimates published in ERDA R&D Report No. 114[2] have been escalated to mid-1975 basis by use of a construction cost index for capital-sensitive estimates, while estimates for catalysts, chemicals, and labor costs have been adjusted using applicable Government Price Indexes.

Population of the complex was originally estimated at approximately 1,700. The original estimate is shown in Table 9 to illustrate the distribution between the modules plus general administrative responsibilities. The operating labor requirements shown in Table 9 were used in conjunction with the symposium economic guidelines to provide the labor component of the operating cost estimate.

D. Required Annual Revenue

The Required Annual Revenue to support a 10% discounted cash flow rate of return (DCF) with 100% equity financing is shown in Table 10. Here the required annual revenue is presented in the form of equivalent uniform annual costs. This method may be used to compare nonuniform time series of money disbursements and receipts at a given discount value.

The present value of each nonuniform disbursement is calculated and then restated in terms of an equivalent uniform annual series. This is a convenient means of showing a single representative cost item when using the DCF method.

Required annual revenue on this basis is approximately $480 million per year. The relative contribution of required revenues, by module, is also shown. Here the relative requirements, in decreasing order of magnitude, are the process area, power plant, and coal mine.

For reference, the required annual revenue adequate to support the coal mine represents approximately 22% of the total. The calculated transfer value of ROM coal is $8.75 per ton. Required revenue to support supply of the clean

Table 9 - Personnel Population Summary

Item	Operating	Maintenance	Administration	Total
Administration	-	-	355	355
Coal Mine	529	125	-	654
Coal Preparation	12	21	-	33
Process	181	300	-	481
Power Plant	101	40	-	141
Offsites	21	26	-	47
Total	844	512	355	1,711

Table 10 - Required Annual Revenue at 10% DCF
and 100% Equity in Equivalent Uniform Annual Costs
($ Million/Year)

Cost Item	Coal Mines	Coal Preparation	Process	Power Plant	Offsite	Total
Total Operating Costs	64.7	3.6	59.0	44.0	13.0	184.3
Income Taxes	11.2	3.3	47.4	41.8	8.9	112.6
Capital Investment						
Fixed Capital Investment	17.4	4.8	69.4	62.6	13.1	167.3
Startup Costs	0.3	0.1	1.1	1.0	.2	2.7
Initial Catalysts & Chemicals	-	-	.8	-	-	.8
Recurring Capital Investment	8.1	-	-	-	-	8.1
Working Capital	2.6	.2	3.4	2.5	.7	9.4
Total Capital Investment	28.4	5.1	74.7	66.1	14.0	188.3
By-Product Credit						
Sulfur	-	-	(7.4)	-	-	(7.4)
Required Annual Revenue	104.3	12.0	173.7	151.9	35.9	477.8
Percent of Total	21.8	2.5	36.4	31.8	7.5	100

feed coal to process unit is approximately 24% of the total revenue; a calculated transfer price for the clean coal to the pyrolysis unit is approximately $14.00 per ton.

E. Required Product Selling Prices

Required product selling prices are depicted in Figure 5. Values are shown for the cases of 100% equity and 65/35 debt-equity basis and represent the combinations of selling prices required to achieve the 10% DCF. The electric power selling price is bus-bar basis and the syncrude value is F.O.B. the plant. This relationship is derived directly from the required annual revenue and production rates for power and syncrude.

For example, with sale of electrical power at 40 mills and 100% equity financing, the required syncrude price would be about $24 per barrel; for 65/35 debt-equity ratio, the syncrude sales price would drop about $18 per barrel. Also shown in Figure 5 is a boundary case, with zero tax burden and 0% DCF. This line is related to a break-even position.

Figure 6 presents the selling prices for the maximum and minimum oil production cases as described in Table 3, 100% equity basis. To illustrate, for the minimum oil production case, sale of electrical power at 40 mills would require sale of syncrude at $16 per barrel.

The required revenue reported here in mid-1975 dollars is 40% greater than that reported in Reference 2 in first quarter-1974 dollars. The major influence on the increased projected required selling prices during this interval was the 30% increase in estimated fixed capital investment.

F. Sensitivity Analysis

The sensitivity of required selling prices to the DCF level is given in Figure 7 where the sensitivity over the range of 0 to 20% DCF is shown. The results illustrate that the required product prices are highly sensitive to the DCF; increase in required DCF to 15% for the 100% equity case would increase the 201
required selling price by about 30%; for the 65% debt case, the increase is

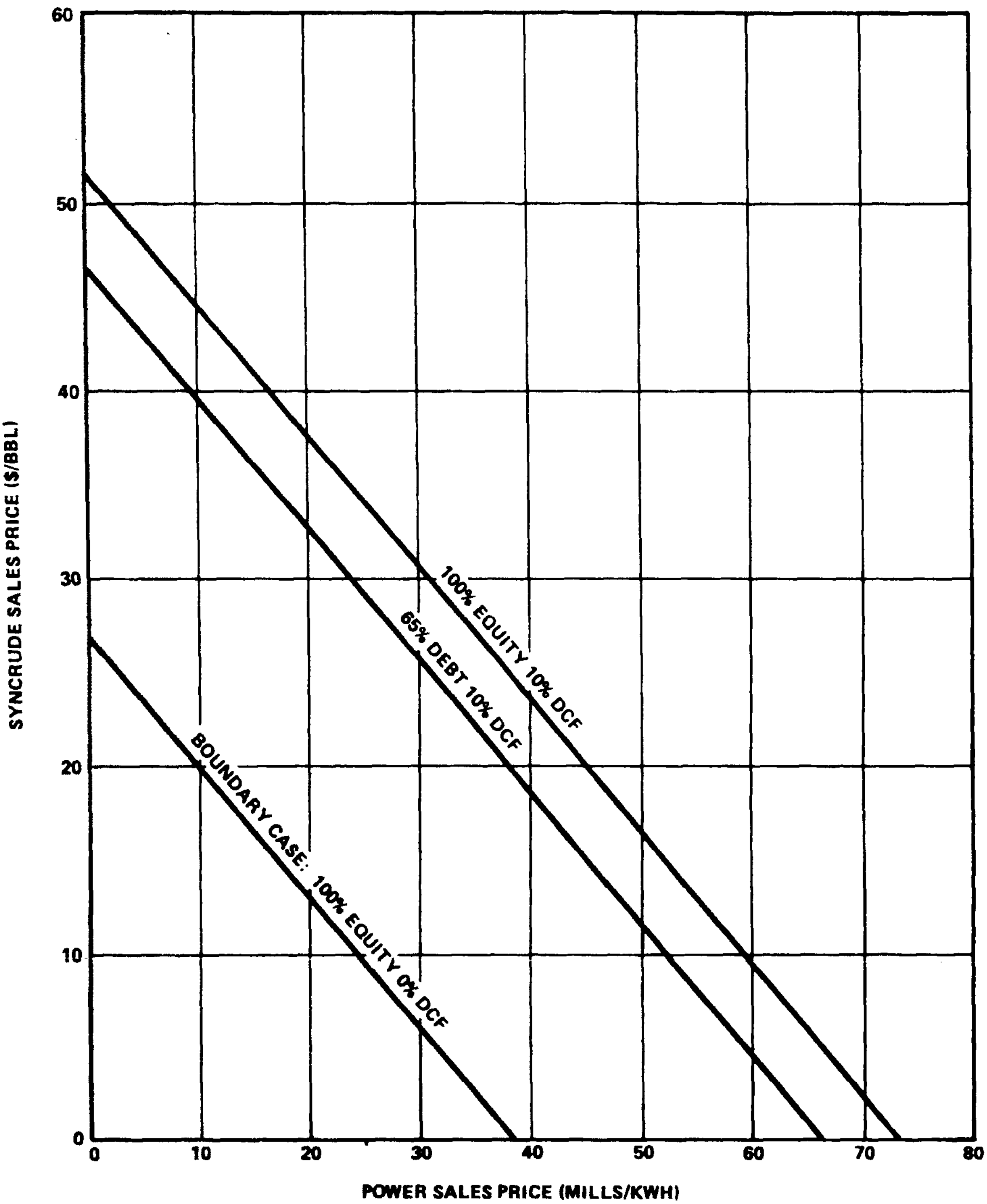


Figure 5 - Required Product Selling Price
Typical Coal Analysis

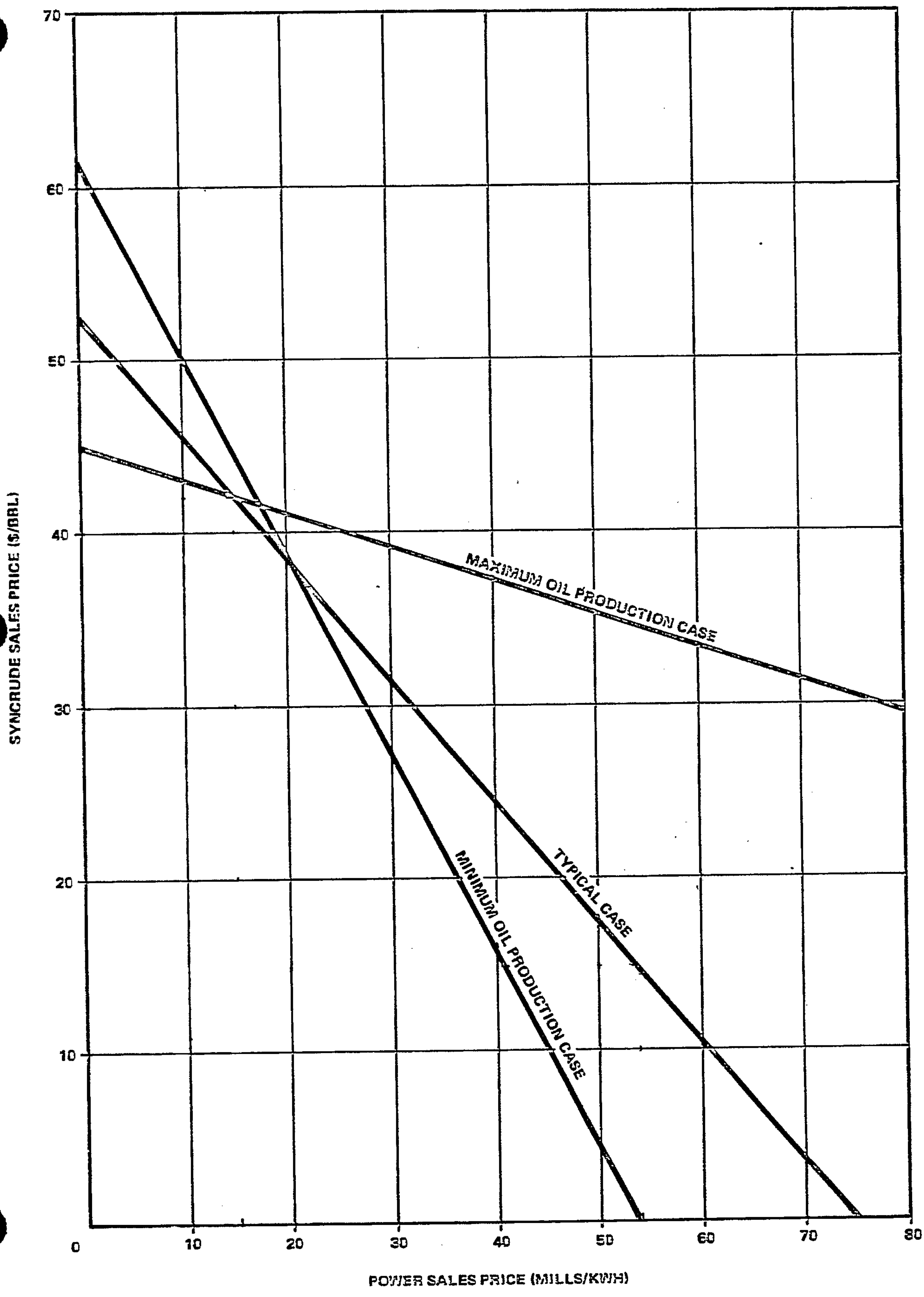


Figure 6 - Required Product Selling Prices for Expected Range of Coal Analysis

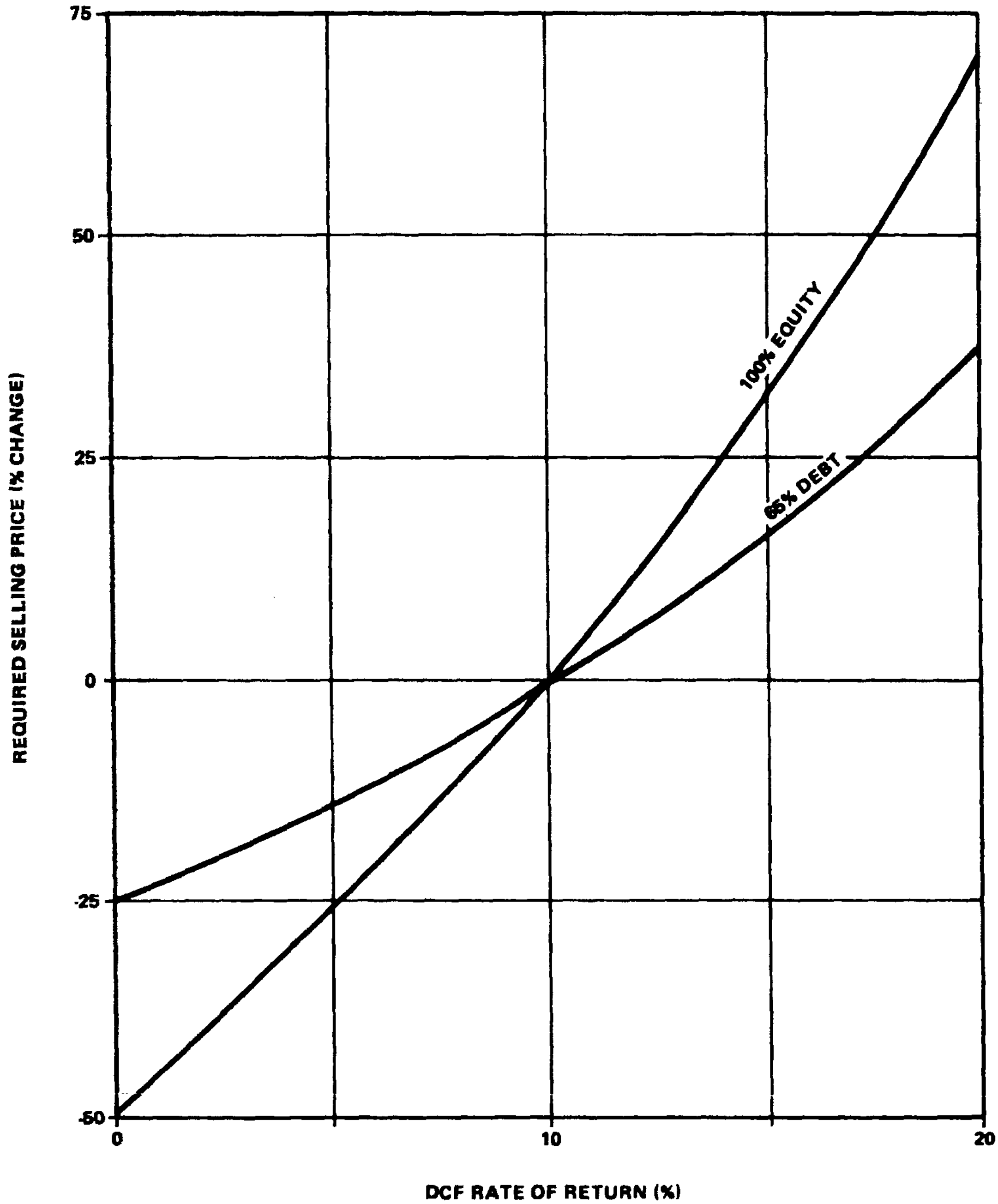


Figure 7 - Sensitivity of Required Annual Revenue to Changes in % DCF

about 15%. Table 11 highlights factors that are affected by changes in the DCF rate of return. The influence of income tax on the results is significant; an increase of DCF from 10 to 20% increases income tax payments by about 140%.

The sensitivity of required selling price to total capital investment is shown in Figure 8. Over the range of capital investment change of -20% to +20%, the sensitivity relationship is linear. The sensitivity factor is 85%; for example, a 10% change in capital investment would result in an 8.5% change in the required product selling price.

The sensitivity of required selling price to operating costs other than capital investment sensitive costs is small. For example, increasing these operating costs by 10% would increase the required product selling price by only about 1%.

G. Comparison with Single Feed Coal Composition (Point) Design

An accurate estimate of the economics of a traditional point design for comparison of the results presented in this paper would require considerable additional effort and is beyond the scope of the work. However, a guidance type estimate has been developed as orientation to indicate the probable economic impact of designing for a range of coal characteristics vis-a-vis a point design. This general "guesstimate" of point design economics was obtained by judgment of fixed capital investment and operating costs for a point design in combination with sensitivity curves presented earlier.

A comparison of estimated fixed capital investment in millions of dollars is:

	Feed: Range of Coals	Single Coal Source
Coal Mine	132.5	132.5
Coal Preparation	37.0	37.0
Process	529.0	490.0
Power Plant	477.0	381.0
Offsites	99.5	99.5
Total	1,275.0	1,140.0

Table 11 - Required Annual Revenue at Various DCFs
($ million/year)

Cost Item	Discounted Cash Flow Rate of Return		
	0%	10%	20%
Operating Costs	184.4	184.4	184.4
Depreciation	73.5	73.5	73.5
Income Taxes	---	112.6	278.4
Return on Investment	---	115.1	284.1
By-Product Credit	(7.4)	(7.4)	(7.4)
Required Annual Revenue	250.5	478.0	813.0
Percent of Total	52	100	170

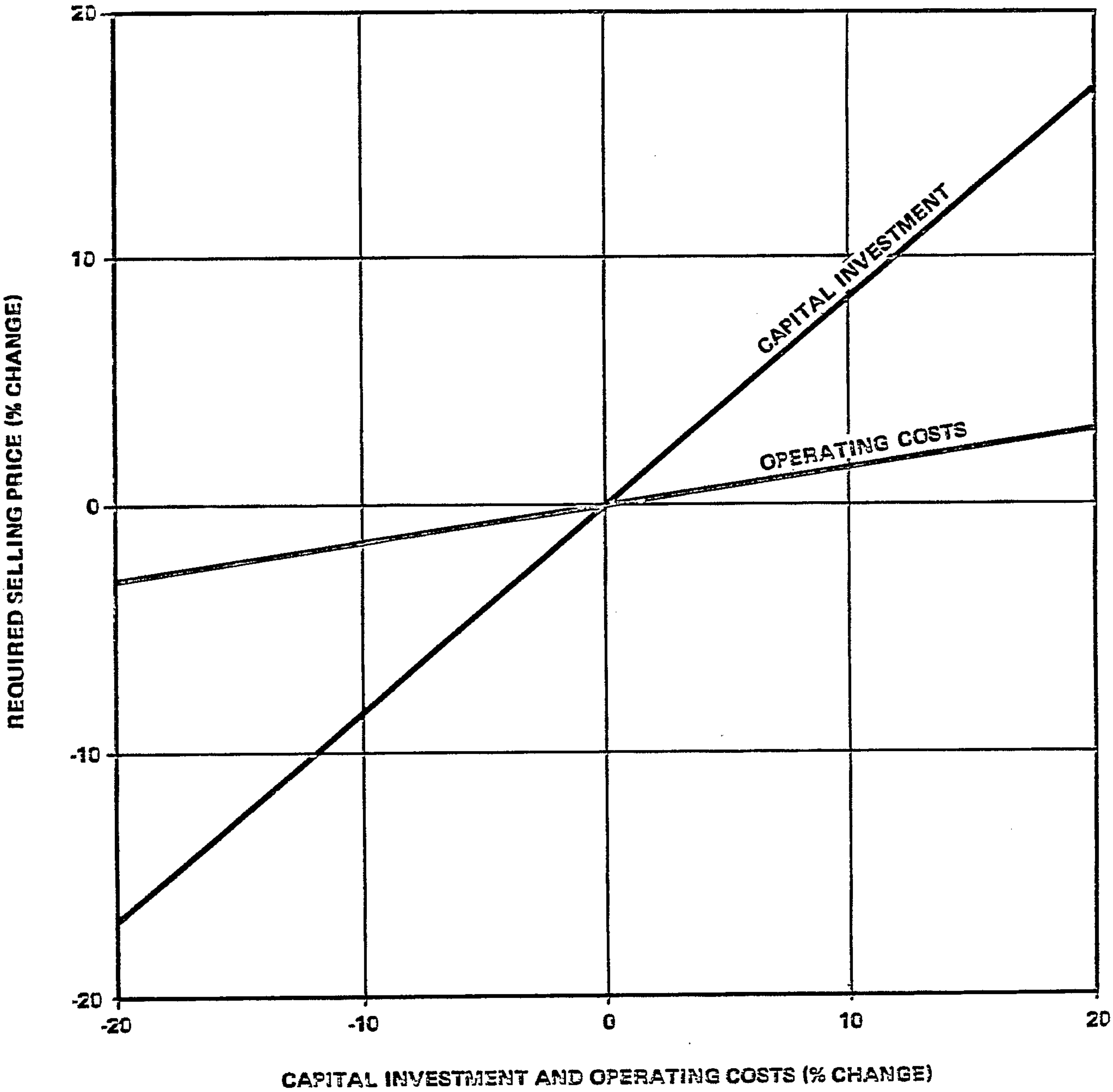


Figure 8 - Sensitivity of Required Selling Price to Changes in Capital Investment and Operating Costs

For operating costs, the effect of coal characteristics was judged to be minor. The major impact is caused by a fixed capital investment requirement.

Using the difference in fixed capital investment in combination with sensitivity curves, we estimate that the 10.6% reduction in the fixed investment would result in an 8 - 9% reduction in required selling prices, to achieve a 10% DCF rate of return on either of the two cases, with 100% on equity or 65% borrowed capital.

VIII. SUMMARY

A conceptual design for a commercial COED-based coal conversion complex to produce syncrude, electrical power, and pure sulfur has been described. The complex would mine approximately 36,000 tons of ROM coal per day and, for a typical feed coal analysis, produce about 28,000 BPD of syncrude, 830 megawatts of electrical power, and 850 TPD of sulfur. The fixed capital investment is estimated to be about $1.3 billion mid-1975 dollars; the total capital requirement is somewhat in excess of $1.4 billion.

For a 10% DCF, typical required selling prices are: when electrical power is sold at 40 mills/kWh, the syncrude would sell at about $24 per barrel. These estimated selling prices have increased 40% relative to the first quarter 1974 values earlier published in an ERDA R&D Report[2]. The change in selling price estimates within a little over a year is sobering.

Required product selling prices are highly sensitive to DCF and capital investment, and only slightly sensitive to operating costs other than those dominated by capital cost.

The design reported here can handle expected variations in feed coal composition. To do so adds fixed capital investment and results in variable product rates. A second-order guidance-type estimate indicates that a single coal feed composition design would result in predicted reductions in fixed capital investments and required product selling prices of about 10% and 8%, respectively. The variable feed composition design is considered the more realistic.

IX. ACKNOWLEDGMENT

We acknowledge the guidance and support provided by ERDA-Fossil Energy, and its predecessor in this effort, Office of Coal Research.

A design and economic evaluation of this type requires the contributions of many people. All such contributions are acknowledged; particular mention is made of Messrs. C.L. Crawford and N.E. Jentz for their process work and Messrs. H.F. Hincks and G.H. Hervey for their project work.

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APPENDIX
ECONOMIC BASIS

Use Mid-Year 1975 $	
Project Life	20 years
Operating Factor	330 days/year
Capital Investment	
Cost of Capital	10%
Working Capital	60-day inventory 60-day cash supply
Land Required	$5000/acre
Startup Expense & Organization	2% of capital investment
Annual Operating Cost	
Feedstock, $/ton	$1.50/ton royalty,(mine included)
Utilities	
Power ($/kWh)	0.015
Water ($/M gal)	0.05
Fuel ($/MM Btu)	1.50
Operating Labor	$15,000/man-year
Operating Labor Supervision	15% of operating labor
Maintenance	
Labor	2% of facilities investment
Supervision	15% of maintenance labor
Materials	2% of facilities investment

Administrative & Support Labor	20% of all other labor
Payroll Extras (fringe benefits, etc.)	20% of all other labor
Insurance	2% of facilities investment
General Administrative Expenses	2% of facilities investment
Taxes (Local, State, & Federal) (No investment tax credit)	50% of net profit
Depreciation	Straight line
Depletion Allowance	10%
By-Product Credit	
Sulfur	$30/ton

COLLECTED WORK NO. 14

Synthetic Fuels from Coal by Fischer-Tropsch

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CHEMICAL ENGINEERING PROGRESS

August 1975

Synthetic Fuels from Coal by Fischer-Tropsch

An economic analysis of the potential for commercial scale operation of this process in the expected future United States fuels market situation.

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The Ralph M. Parsons Co.
Pasadena, Calif.

Preliminary results from a survey being conducted on the potential use of Fischer-Tropsch technology in the United States indicate that a small plant based on the process is not attractive based on current economics, but that a very large facility could offer more interest.

If present barriers to scale-up operations can be overcome, the technology might find application, the interim report shows. A key is the flexibility of the technology. Another major point is the fact that liquid product has no sulfur and a solid particulate content. The product, therefore, should be valuable if used as a blend stock to achieve the maximum sulfur and solids concentrations allowed to satisfy environmental standards.

Fischer-Tropsch technology has been practiced on an industrial scale in several coal-based economies. (*1, 2, 3*) However, it has never been practiced successfully in an economy where its products faced open competition from indigenous or unrestricted low-tariff importation of crude oil.

The U.S. is involved in a major effort to develop a viable technology for conversion of indigenous high-sulfur coal to ecologically "clean" liquid, gaseous and solid fuels. The Ralph M. Parsons Co. is playing a role in that program by supplying technical evaluation and preliminary design services to the Office of Coal Research, now a part of the Energy Research and Development Administration (ERDA). During its work, Parsons has twice been asked to make preliminary investigations of Fischer-Tropsch technology. This article summarizes the results of investigative work performed to date, and briefly discusses future plans.

The objective of the work was to investigate the question of whether the Fischer-Tropsch technology should have a place in future U.S. synthetic fuel production plans—and, if affirmative, to provide guidance and concepts for a program designed to define the Fischer-Tropsch role.

Work to date has consisted of completion of two separate tasks. In the first case, a preliminary design and economic evaluation was developed based on production of synthesis gas (syngas) in a single large high-capacity shop-fabricated gasifier. The primary product objective was fuel oil, and the secondary objective was substitute natural gas (SNG). The plant would produce approximately 2,500 bbl./day of fuel oil and approximately 60-million std.cu.ft./day of SNG.

The second case was a conceptual design prepared under short deadline pressure during the "Project Independence Blueprint" effort. Fischer-Tropsch technology was not included in the first round of assessment of coal conversion processes. During review of the options, the reviewing committee decided to include a large Fischer-Tropsch plant. Parsons was invited to prepare within a ten-day period a conceptual description of such a facility and its projected economics. This was done. The guidelines included facility production of 100,000 bbl./day of light fuel oil plus significant amounts of SNG; preliminary analysis resulted in production of approximately 1,700-million std.cu.ft./day of SNG.

The nucleus of the following presentation therefore spans what could be the extremes of a spectrum of possible Fischer-Tropsch applications, specific cases being: 1) a small Fischer-Tropsch plant, emphasizing fuel oil plus SNG production; and 2) a large Fischer-Tropsch-based complex, emphasizing light fuel oil plus SNG. Each of the two cases will be briefly summarized, followed by a discussion of the results.

Small plant would process 7,130 ton/day coal

The small plant is conceived to be a grass-roots manufacturing complex located in Appalachia. Its characteristics, with nameplate capacity to process 7,130 ton/day of run-of-mine coal, are described in the following.

The plant will produce the following:

1. SNG with a higher heating value of 960-Btu./std.cu.ft., at a pressure of 1,000 lb./sq.in.
2. Fuel oil with nil sulfur content and an expected boiling range of pentanes through 850°F. Its heating value is estimated to be 5.4 million Btu./bbl. (20,750 Btu./lb.) and its specific gravity 58.4°API.
3. Sulfur; 99.5% minimum purity.
4. Slag.

Table 1. Typical analyses, wt.-%

Proximate analysis:	
Moisture, as received	2.1
Volatile matter, MAF	37.0
Fixed carbon, MAF	53.6
Ash, MAF	7.6
Heating value, Btu./lb.	
Gross	13,317
Net	12,812
Ultimate analysis (as received):	
Carbon	75.0
Hydrogen	5.2
Oxygen	8.7
Nitrogen	1.2
Sulfur	2.6

Fuel gas and waxes are also produced and are used as in-plant fuel.

Raw materials will consist of: 1) run-of-mine Pittsburgh No. 8 seam coal, with the typical analyses as shown in Table 1; 2) oxygen, 99.5%, produced in the plant; and 3) raw water. The run-of-mine coal is cleaned, washed, and sized in preparation for feed to the process units.

The coal conversion process is based on gasification of the process coal followed by a catalytic synthesis of liquids from the carbon monoxide and hydrogen mixture produced during gasification. The bulk of the material not liquefied is converted to SNG. The process is depicted in the block flow diagram shown in Figure 1.

The basic design criteria for the gasifier vessel specified that it be a single unit of near-maximum commercial size, shop-fabricated, and transportable by rail or waterways to the Appalachian region. On the basis of maximum weight and dimensions of the vessel operating at about 1,050 lb./sq.in., the capacity was set at 3,500 ton/day of coal. The gasifier is an entrained-bed gasifier, which is a modification of the "Bi-Gas" design under development in the ERDA/Fossil Energy Program. (*4*)

Fischer-Tropsch synthesis is performed in a fixed-bed catalytic converter operating at about 70% conversion. (*5, 6*) Carbon monoxide and hydrogen not converted in the synthesis reactor are admixed with methane produced in the gasifier units, and further processed by shifting and

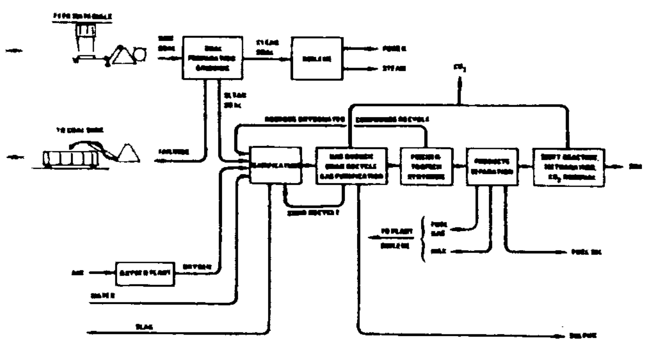

Figure 1. Fischer-Tropsch oil/gas plant process block flow diagram.

methanation to produce SNG.

Water-soluble oxygenated compounds produced in the Fischer-Tropsch reaction are separated and recycled to extinction by returning this stream to the gasifier. (*7*)

Major sections of process facility

For convenience, the production of clean fuels from coal in the Fischer-Tropsch plant can be divided into the following areas: coal mining and preparation; coal gasification; raw gas quenching and purification; Fischer-Tropsch synthesis and products separation; catalytic conversion of the tail gas to produce additional methane; and a cleanup section for the final removal of residual CO_2 and water from the SNG. Several key elements will be described.

Coal preparation and grinding. Coal from a 30-day supply stockpile is conveyed to a washing and primary crushing plant. Refuse from the plant is returned to the mine area for disposal. About 4,630 ton/day of clean coal are produced and subsequently passed to a pair of grinding mills to reduce the size of the coal particles so that 70% will pass through a 200-mesh screen.

A pyritic sulfur removal process, such as is being developed by TRW Inc. under the sponsorship of the Environmental Protection Agency, (*8*) reduces total sulfur content of this steam coal from 2.6 to 0.8 wt.-%.

To prepare the feed to the gasifier, 3,500 ton/day of wet ground coal are transferred to a coal slurry tank. Recycled Fischer-Tropsch process water and makeup water are added to the coal to make a suitable coal/water mixture for pumping.

Gasification. High-pressure entrained gasification was selected to achieve an attractive capacity for a given size reactor, to facilitate removal of acid gases, to increase the production of methane, and to obtain a favorable hydrogen-to-carbon-monoxide ratio.

Fischer-Tropsch synthesis. Purified syngas combined with recycled tailgas flows to the Fischer-Tropsch fixed-bed catalytic reactors. (*6*) Heat released by the exothermic reactions is absorbed by boiling water on the shell side.

Most effluent is cooled further in heat exchangers and flashed in a cold separator. Part of the gas is recycled to the reactor inlet; the rest passes through a refrigerated absorber system for recovery of light hydrocarbons and gasoline. Cleaned tailgas is then passed to the shift reaction system. The aqueous phase from the separator—after alcohol recovery—is recycled to extinction in the gasifier. The hydrocarbon phase is sent to a stabilizer to make storable fuel oil and off-gas, which is used as in-plant fuel.

Shift, methanation, and final cleanup. Clean tailgas from the Fischer-Tropsch reaction is subject to shift conversion whereby carbon monoxide reacts with steam to produce hydrogen and carbon dioxide. Gas from the shift conversion is sent to a methanation unit for conversion of residual carbon monoxide to methane. The final steps are: 1) cleanup of carbon dioxide in an acid gas removal unit, and 2) drying of the SNG in a silica gel dryer.

Predicted compositions of raw gasifier product, the feed

Table 5. Fischer-Tropsch economic estimating basis, small plant

1. All estimates in 1st quarter 1975 dollars.
2. Location: Appalachia.
3. Working capital; $15-million, including an allowance for spare parts inventory.
4. Facility construction cost: $205-million.
5. Cost of initial catalysts and chemicals: $1-million.
6. Project design and construction period: 4 years.
7. Project life: 20 years of operation.
8. Coperating rate: 330 stream days per year.
9. Production rate in initial operating years: 75% for first year, and 100% for second year.
10. Income tax: 48% federal; 3% state; 7% investment tax credit on process plant facilities.

unusual foundation conditions; process licensing fees; and escalation.

Profitability analyses were developed to provide preliminary assessment of profitability potential for the Fischer-Tropsch technology. Key bases used for the estimate are summarized in Table 5.

Production costs were estimated at $30.6 million/yr., equal to $1.35/million Btu. The selling price required to achieve a 12% discounted cash flow rate of return after tax was calculated at $3.55/million Btu., with 100% equity financing. With a 75/25 debt to equity ratio at a 9% interest rate, the required selling price is calculated at $3.00/million Btu.

Study of large Fischer-Tropsch plant

The second survey, performed under deadline pressure, was for a conceptual plant to explore the potential for a future large facility to produce 100,000 bbl./day of fuel oil plus a significant amount of SNG. We conducted a brief summary of scope and objectives for this ten-day judgmental analysis, which provided a vehicle for informed conceptualization of a plant using procedures not proven in practice. The scope and objectives are shown in Table 6.

The process scheme was similar but not identical to that used for the small plant, shown earlier in Figure 1. Changes included: 1) washed, sized coal was purchased; 2) electricity was purchased; and 3) field-fabricated vessels and reactors, with diameters to 25 ft., were used.

Overall material balance is shown in Figure 3. Here, we see that 140,000 ton/day of coal are used to produce 100,000 bbl./day of fuel oil plus approximately 1,700 million std.cu.ft./day of SNG and somewhat less than 5,000 ton/day of sulfur. Oxygen and water requirements are approximately 50,000 and 55,000 ton/day, respectively.

Characteristics of this process scheme include production of a significant amount of methane in the entrainment-type gasifier, operated at approximately 1,000 lb./sq.in.gauge plus a single pass through the Fischer-Tropsch synthesis reactor. This is followed by recovery and purification of liquid products and conversion of gas stream to SNG by shift reaction, methanation, and carbon dioxide removal. Approximately 21% of the methane that appears in the SNG is produced in the methanation step.

Using the parameters previously described for process and plant, our judgmental estimate is that the FCI for the 100,000-bbl./day Fischer-Tropsch plant might be $2.8-billion, based on first quarter 1975 cost. Economic parameters are summarized in Table 7.

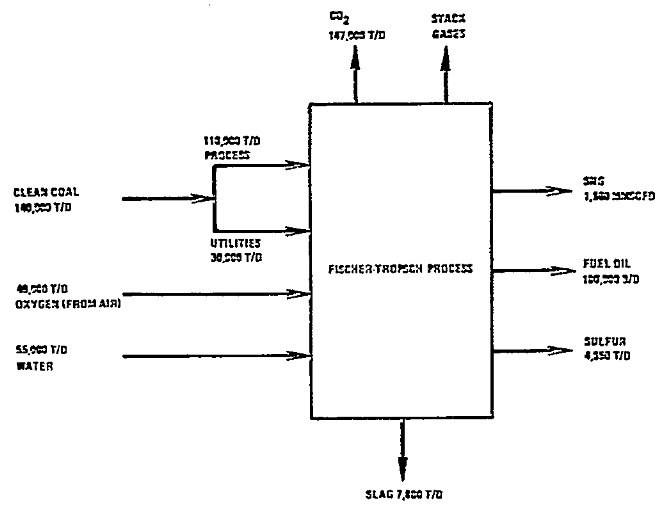

Figure 3. Fischer-Tropsch overall material balance, large plant.

Table 6. Fischer-Tropsch statement of objectives, large plant

1. Develop rapid planning-type judgemental definitions of characteristics and approximate economic input requirements for a liquid-producing coal conversion facility to produce 100,000 barrels per day of liquids plus significant SNG.
2. Develop rapid estimate of the allocation of resources required to construct and operate the facility.
3. Develop rapid estimate of development, design, procurement, and construction schedules for a U.S.A. commercialization of the technology.
4. To consider and define the national restraints that would limit the ability to successfully complete the project.

Table 7. Fischer-Tropsch economic estimating basis, large plant

1. All values in 1st quarter 1975 dollars.
2. Interest rate: 9%.
3. Commitment fee on construction load: 0.5%.
4. Debit/equity ratio: 75/25.
5. Project life: 20 years.
6. Depreciation schedule: 20 yr., straight line.
7. Design/construction schedule: 5 years.
8. Working capital: $200-million.
9. Startup costs: $225-million.
10. 330 stream days per calendar year.
11. Coal price: $10.00/ton, and $15.00/ton.
12. Discounted rate of return (DCF): 12%.

Table 2. Fischer-Tropsch gas composition summary, small plant

Component	Gasifier raw gas	F-T syngas feed	SNG
	Composition in vol.-%		
H_2	22.1	44.2	2.3
CO	15.9	31.8	0.1
CO_2	17.2	1.0	0.9
N_2	0.4	0.7	2.3
CH_4	11.0	22.1	94.4
H_2O	32.8	0.1	0.008
H_2S	0.6	0.0001	–

Table 3. Fischer-Tropsch summary of gas stream characteristics, small plant

Factor	Gasifier raw gas	F-T syngas feed	SNG
	Stream description		
H_2/CO ratio	1.39	1.39	–
H_2 – CO, vol.-%	38.0	76.0	2.4
Mol weight	20.42	14.01	16.7
Dry gas flow rate, million std.cu.ft./day	247.0	183.8	58.0
Dry gas HHV, Btu./std.cu.ft.	349	–	961
Dry gas LHV, Btu./std.cu.ft.	316	–	865

to the Fischer-Tropsch unit, and the product SNG are shown in Table 2. Characteristics of these streams are shown in Table 3; factors given include the hydrogen-to-carbon monoxide ratio, flow rate, and heating value.

The overall material balance is depicted in Figure 2.

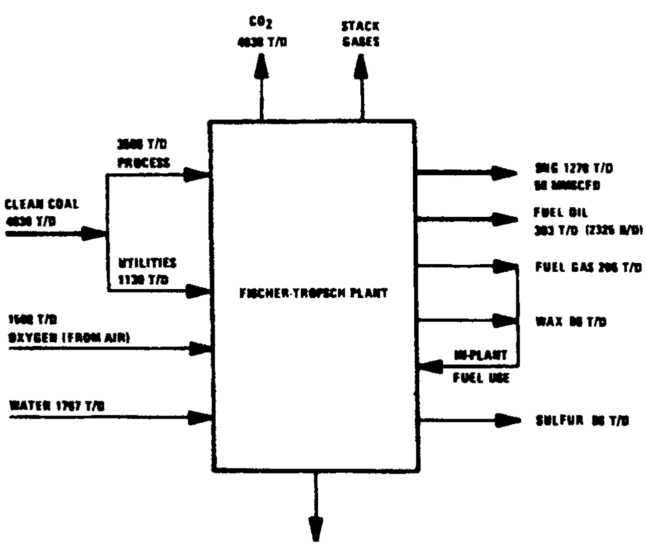

Figure 2. Fischer-Tropsch overall energy balance, small plant.

Most of the necessary energy required for the operation of the complex is derived from the 1,130 ton/day of coal fed to the utility boilers. Fuel gas and wax produced by the Fischer-Tropsch process supplement in-plant fuel needs.

The stream day production rates of salable products are estimated to be 58 million std.cu.ft. of SNG, 2,325 barrels of fuel oil, and 88 tons of sulfur.

$205-million for capital investment

A preliminary factor-type fixed capital investment was developed for a grass-roots process complex, consisting of the process plant previously described, plus necessary ancillary facilities to support the plant and plant population. The costs of the coal mine and coal preparation plant were estimated based on defined equipment and support facility requirements.

Estimated fixed capital investment (FCI) for construction of this complex totals $205-million including construction, engineering, and sales tax. All costs are based on first-quarter 1975 dollars. A summary of the estimate is shown in Table 4. Mineral rights cost of $1.00/ton of clean washed coal payable as mined and processed were used.

In addition to the $205-million FCI for the mine and process plant facilities, it is estimated that the project will require the following additional capital expenditures: plant startup costs, $10-million; land for process plant, $500,000; initial catalyst and chemicals, $1-million; for a total of $11.5 million, and a grand total of $216.5 million.

Not included in the FCI are owner's expenses; land acquisition, water, mineral and process rights, and rights-of-way; taxes other than sales and payroll taxes; working capital, interest, and financing; raw materials and initial supplies; client's permits; premium time costs; piling or

Table 4. Fischer-Tropsch preliminary fixed capital investment summary, small plant

Unit description	Construction cost $1,000's
Coal mine, coal preparation, and coal wet grinding	18,100
Coal slurry feed system, coal gasification, raw gas quench, and char separation	24,900
Raw gas purification	23,800
Sulfur plant	5,600
Fischer-Tropsch synthesis, catalyst preparation, and F-T products separation	27,300
Shift conversion, and methanation	7,600
CO_2 removal and drying	4,900
Oxygen plant	16,800
Slag disposal	400
Coal sulfur removal unit	7,000
Utilities, buildings, offsite storage, water and waste treating, site preparation, roads and paving	39,100
Total construction costs	175,500
Home office engineering, and sales tax	29,500
TOTAL	205,000

Projected production costs, estimated for delivered run-of-mine coal costs at $10.00 and $15.00/ton, are as follows: $1.02/million Btu. for $10.00/ton coal; and $1.35/million Btu. for $15.00/ton coal.

The selling price required to achieve a 12% discounted cash flow rate of return was calculated for two financing methods. With 100% equity financing, the required selling price is $2.21/million Btu. for $10/ton coal and $2.53 for $15/ton coal. Using a 75/25 debt-to-equity ratio with interest at 9%, the selling price is $1.60 using $10/ton coal and $1.95 using $15/ton coal.

Summary

This is an interim report on a survey of the potential use of Fischer-Tropsch technology in the U.S. However, certain intermediate conclusions are drawn and presented. These interim conclusions will be used as a base for future work in the field. The Ralph M. Parsons Co. gratefully acknowledges the support and guidance of ERDA/Fossil Energy for the work described here. #

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A CEP CAPSULE

COAL PROCESSING:

Fischer-Tropsch Plant Design Criteria

When completed, this multi-product complex will process 30,000 ton/day of coal into nearly 50,000 bbl./day of liquid fuels and 250 million std. cu. ft./day of SNG.

J. B. O'Hara, A. Bela, N. E. Jentz, and S. K. Khaderi
The Ralph M. Parsons Co., Pasadena, Calif.

Coal-based Fischer-Tropsch plants were successfully operated in Germany during World War II, and in South Africa for the past 20 years. The process has been studied in the U.S. by the Bureau of Mines, now part of the Energy Research and Development Administration–Fossil Energy.

In 1973, The Ralph M. Parsons Co. considered the question: Is there a place for Fischer-Tropsch technology in future U.S. synthetic fuels production plans? And if so, what role might it play? To answer this, Parsons completed a conceptual design and economic evaluation for a small plant to process 4,600 ton/day of coal and produce about 2,500 bbl./day of liquid fuels plus about 60 million std. cu. ft./day of substitute natural gas (SNG). Subsequently, Parsons developed a possible configuration and approximate economics of a large Fischer-Tropsch complex to produce 100,000 bbl./day of liquid fuels plus significant SNG for the Project Independence effort. The results of these two preliminary assessment efforts were published last year. (*1*)

The design/economics now being developed represents a comprehensive follow-up effort to describe a facility to produce synthetic fuels from coal to be responsive to future U.S. demands. The concept is that the complex should be large, simple, and produce both liquid fuels and SNG. It is important to note that liquid fuels produced by Fischer-Tropsch technology are premium fuels; they do not contain sulfur, nitrogen, or solid particulates. The conceptual design discussed here is conceived for operation in the 1980s.

Parsons is actively assisting ERDA to develop viable commercial plants for the conversion of coal to clean fuels. This role is comprised of two distinct elements:

1. Preliminary design services in which Parsons develops preliminary conceptual designs and economic evaluations for commercial coal conversion plants.

2. Parsons also supplies technical evaluation contractor services to assist ERDA in monitoring certain of the liquefaction development programs.

Principal elements of a multi-product Fischer-Tropsch complex are illustrated in Figure 1. The complex will contain the following units: a coal mine, a coal preparation plant, a steam-oxygen coal gasification plant, gas purification facilities, sulfur production facilities, a Fischer-Tropsch synthesis unit, a liquid products recovery/separation section, as well as SNG production, utilities production and distribution, ancillaries, and support facilities.

Predesign studies

Predesign study results provided the basis for defining the preferred design criteria before beginning detailed design work. Major gasification technologies were reviewed and analyzed. The plan is to use slagging, medium pressure, entrainment gasification for this conceptual design.

Candidate synthesis reactors were reviewed. A key objective was to define potential high capacity converters; a preliminary economic comparison between the alternative reactor systems is in progress. Based on available analysis results, the plan is to use a design with the catalyst sprayed on heat transfer surfaces. This concept is based on work done by the U.S. Bureau of Mines. (*2*) Particular attention is being paid to design factors required to achieve high unit capacity, such as kinetic and heat transfer factors.

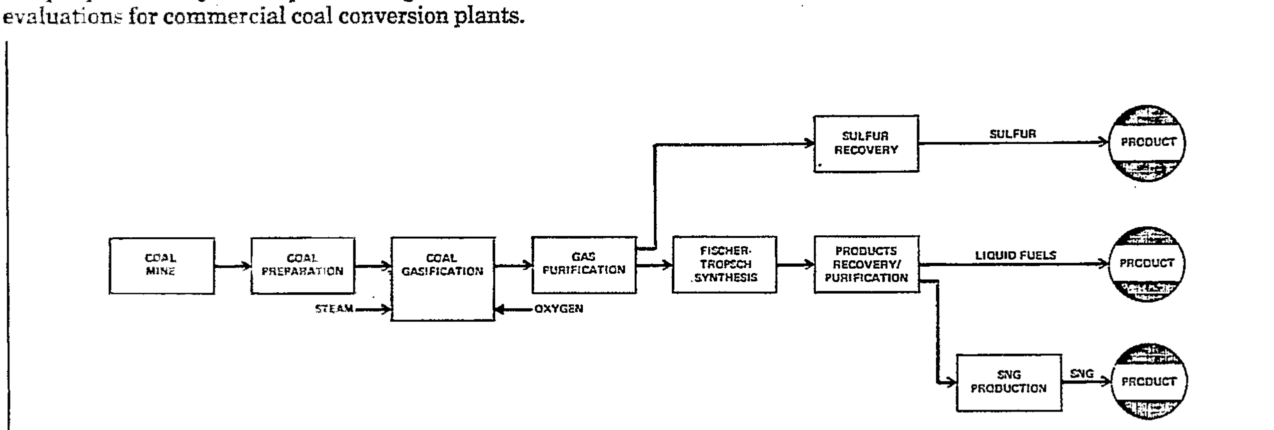

Figure 1. Simplified block flow diagram of the Fischer-Tropsch process.

During an extensive effort to maximize thermal efficiency of the complex, all significant contributors to energy efficiency improvement were analyzed. Our present opinion is that thermal efficiency for the conceptual commercial complex will be of the order of 70%, with this efficiency expressed as the ratio of B.t.u. content of fuel products divided by the B.t.u. content of the feed coal, × 100. This is a significant improvement over past design results.

Design criteria are intended to 1) describe key elements of the design that will permit users to anticipate size, product slate, and general characteristics of the resulting facility, and 2) permit designers to proceed with their objectives and work.

Design parameters provide a capacity of 500 billion B.t.u./day in liquid and gas products (the plant will be comprised of two parallel trains of 250 billion B.t.u./day each). The site location will be the Eastern region of the U.S. Interior (coal) Province, and the feed will be Illinois No. 6 seam coal produced in a captive mine. It will be a grass roots complex complete with all ancillaries required to support the facility and its operation (all utilities will be captively produced). Liquid products will consist of LPG, naphtha, diesel fuel, premium fuel oil, and alcohols. Product SNG will be of pipeline quality and byproduct sulfur will also be produced, Figure 2.

Probable configuration

Coal mine: approximately 40,000 ton/stream day of run-of-mine coal will be produced in four separate mining units.

Coal preparation: a series of jigs, screens, cyclones, centrifuges, and roll crushers wash and reduce the particle size of the coal. The stream-day demand for the gasification plants is about 30,000 tons.

Gasification: the coal is fed to intermediate pressure entrainment, slagging-type steam-oxygen gasifiers. The product gases will pass through a steam superheater and a waste heat boiler, and the contained char will be removed.

Shift conversion and acid gas removal: approximately one-third of the syngas is subjected to sour shift conversion, and the shift gas joins the feed stream to an acid gas removal plant. The carbon dioxide effluent stream is exhausted to the atmosphere and a hydrogen sulfide-carbon dioxide mixture is fed to a sulfur production plant.

Synthesis reactors: the sulfur free synthesis gas passes to the synthesis reactors where liquefaction occurs and steam is generated to maintain proper control of reaction temperature.

Products recovery/purification: the synthesis reactor effluent gas mixture is separated from the liquid products, which are fractionated into naphtha, diesel fuel, premium fuel oil, and alcohols. Acids formed are separated and processed for ecologically acceptable disposal.

Methanation: the gas stream from product recovery is fed to the methanation section where SNG is produced. The SNG is dried and compressed to a pressure of 1,000 lb./sq. in. for delivery to a pipeline.

Waste heat recoveries: waste heat will be recovered throughout the process as steam, which will be used as process feed in turbines for prime mover drives, and for electrical power drives. Steam is generated at a pressure of 1,250 lb./sq. in.

A report describing the design and projected economics of the process is to be issued following completion of the work. The reliability of the Fischer-Tropsch technology has been proved by industrial operations. It is a candidate for the U.S. synthetic fuels programs.

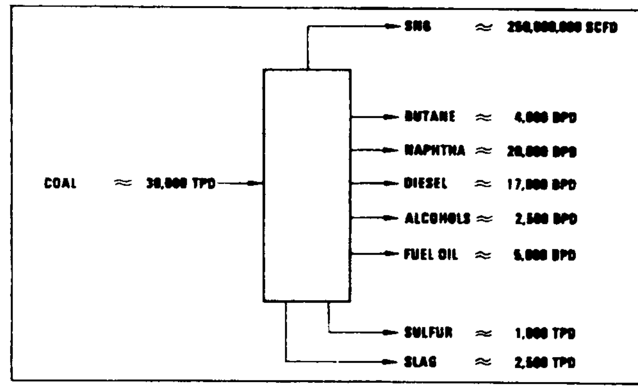

Figure 2. Expected product output of the Fischer-Tropsch process.

Acknowledgment

Many people and organizations contribute to a broad scope design and economic evaluation of this type. The contributions and support of ERDA-Fossil Energy are gratefully acknowledged. Many people in Parsons have added their contributions. Those of Mr. D. G. Reynolds on the process design and Mr. R. D. Howell on the gasification section are particularly noted. #

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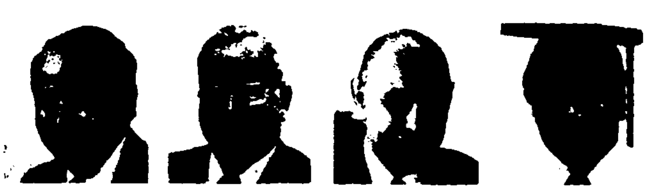

O'Hara Bela Jentz Khaderi

COLLECTED WORK NO. 16

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FISCHER-TROPSCH COMPLEX CONCEPTUAL DESIGN/ECONOMIC ANALYSIS

OIL AND SNG PRODUCTION

R & D REPORT NO. 114 - INTERIM REPORT NO. 3

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Prepared for
ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION
WASHINGTON, D. C. 20545

SECTION 1

INTRODUCTION

This report describes the results of a conceptual design and economic evaluation for a conceptual Fischer-Tropsch plant responsive to U.S. demands and economic requirements.

A primary objective of this conceptual design is to define the characteristics and projected economics of a commercial coal mining and conversion complex to be constructed and operated in the 1980's and 1990's. Key target characteristics of the design include:

- Large size, simplicity, and reliability
- Energy efficiency
- Where justified, incorporation of advanced concepts now in development to achieve stated objectives
- Definition of incentives for further development work required to convert the concepts to reality.

Fischer-Tropsch technology provides potential for broad product flexibility. A range of product spectrums can be produced by proper selection of catalyst, reactor configuration, and operating conditions such as feed gas composition, temperature, pressure, and space velocity.

It is important to recognize that the design presented here represents only one of a large number that can be developed to exploit Fischer-Tropsch technology.

1.1 OBJECTIVES

Objectives of the work described in this report include:

- Develop a conceptual design for a commercial grass-roots coal con version complex based on Fischer-Tropsch technology. The complex is to be responsive to U.S. requirements. It is to include facilities required to:
  - mine coal
  - clean and wash coal
  - convert coal to ecologically-acceptable, premium liquid and gaseous fuels

- Produce fuels at a price competitive with alternate sources
- Develop projected economics for the complex to include the project and financial parameters for design, engineering, procurement, construction, and start-up
- Recommend development work required to assure successful commercial performance of the complex.

1.2 REPORT ORGANIZATION

A summary of key elements in this report is presented in Section 2 to aid in rapid assimilation of the contents.

Sections 3 through 6 present key technical elements of the design. Design parameters and design bases used are summarized in Section 3. Section 4 describes project scope and major units included in the complex. Here major plant units and material flows are depicted in the form of a block flow diagram. A plot plan and artist's rendition of the plant complex are also presented. Section 5 contains detailed descriptions of the separate units that make up the complex. The detailed process flow diagrams with material balances are presented in Section 6.

Sections 7 through 10 summarize key product characteristics and energy-utilization factors for the design. Section 7 presents projected marketability and characteristics of products of the complex. The material balance for the complex is depicted in Section 8. Overall energy balance is presented in Section 9. The utility summary, by unit, is given in Section 10.

Important environmental factors are summarized in Section 11. Facilities included to ensure that effluent streams are properly treated to meet environmental standards are described here. Section 12 presents a summary of plant start-up procedures, recognizing that during normal operation steam requirements for the complex will be generated by process heat recovery facilities.

The list of major equipment size and materials is presented in Section 13. This equipment list, combined with design information previously summarized in the report, provides the basis for the estimate of fixed capital investment. A detailed projected economic assessment is given in Section 14, including capital investment requirements, discounted cash flow (DCF) rate of return printout and key economic sensitivity factors.

The remainder of the report presents supporting data, analyses, and recommendations for future development effort to ensure that the plant will perform as projected.

SECTION 2

SUMMARY

A conceptual design and economic evaluation has been completed for a project to design, engineer, procure, construct, start up, and operate an industrial complex which will mine high-sulfur coal and convert it to a nil sulfur product mix using Fischer-Tropsch technology. The objective was that the complex should be responsive to future U.S. energy requirements and be competitive with alternate energy sources. The results are summarized in this report.

The design basis was developed in cooperation with representatives of ERDA and the work was done with their guidance and support.

As conceived, the complex is located in the Eastern Region of the U.S. Interior Coal Province, which includes portions of Illinois, Indiana and Kentucky. It will mine approximately 40,000 TPD of run-of-mine (ROM) coal from which it will produce about 30,000 TPD of clean, sized coal as feed to the Fischer-Tropsch plant. Here the coal will be gasified, the gases purified, and then reacted to produce liquid products plus substitute natural gas (SNG). The products will be separated and refined ready for sale. Plant products will have an energy value of approximately 525 billion Btu/day, which is about twice the energy value of commercial coal gasification plants planned for construction in the U.S. The plant will consist of two production lines. The plant is designed to meet environmental standards. It should be noted that the design is one of many that can be developed using Fischer-Tropsch technology.

Products from the plant include about 260 MMSCFD of SNG and approximately 50,000 BPD of liquid products. The liquids consist of LPGs, light and heavy naphthas, diesel fuel, fuel oil, and oxygenates (consisting primarily of alcohols). All petroleum liquids produced contain nil sulfur, nitrogen and particulate matter and can be referred to as premium fuels.

Estimated time needed to design, procure, construct and start up the facility is 57 months. The estimated fixed capital investment is approximately $1.5 billion; all economics have been based on fourth quarter 1975 dollars. The total capital investment required is estimated to be about $1.75 billion. In addition to fixed capital requirements, this total includes the cost of initial raw materials, catalysts and chemicals, working capital, allowance for startup costs, and allowance for land acquisition. The cost of financing during design and construction depends on the method of financing, and was added to the $1.75 billion for the separate project cases reported.

The fixed capital investment estimate was independently evaluated by the U.S. Army Engineer Division, Huntsville, Alabama (USAEDH). This work was done under contract to ERDA, Contract No. EX-76-C-01-1759. The USAEDH estimate was approximately 10% lower than Parsons, and they report an indicated overall estimate confidence factor of $\pm 10\%$.

Annual operating costs for the complex are predicted to be about $190 million. Plant population is approximately 2100 people.

Predicted required product selling prices, expressed as dollars per million Btu, for a 12% DCF rate of return and a twenty-year project operating life are:

FINANCING METHOD		
100% Equity	$\frac{\text{Debt}}{\text{Equity}}$ Ratio = 65/35	Break-Even
3.25	2.50	1.45

These values correspond to about $14.80 and $19.40 per barrel equivalent for the 65/35 Debt/Equity (D/E) ratio and 100% equity cases, respectively, based on a heating value of 6 million Btu per barrel. Full details of the economic analysis, including complete sensitivity analyses, are presented.

PROCESS AND PLANT FACTORS

Key characteristics of the complex include:

- Large captive coal mine.
- Use of high capacity gasifiers - each gasifier vessel projected to produce 250+ million Btu/day of energy products.
- Fischer-Tropsch converter design that permits high throughput and recovery of reaction heat at 1,200 pound per square inch steam.
- Design for high thermal efficiency. Predicted thermal efficiency is approximately 70%, expressed as Btu's in salable products divided by Btu's in feed coal, times 100. Predicted efficiency is the result of considerable technical and economic analysis of alternates. Results of these analyses are reported.

The Fischer-Tropsch converter design is based on application of flame-sprayed catalyst (FSC) techniques which have been demonstrated experimentally by what is now the Pittsburgh Energy Research Center (PERC) of ERDA. Similar reactor designs were used for the shift and methanation reaction sections. This type of reactor is projected to provide efficient recovery of reaction heat as steam at a pressure of 1,200 pounds per square inch. As a result, all steam required to operate the plant, produce the necessary captive power requirements, and also produce excess power for sale is generated in the process sections;

a fuel-fired utility plant is not required for normal operation. All utilities are internally generated, i.e., feeds to the process plant consist of coal, air, and water.

This design is intended to aid in defining the potential for large, second-generation coal conversion plants. It incorporates a number of concepts and equipment items that careful analyses indicate have potential advantages and good probability for high performance. A number of these items is based on commercialization of expected favorable results of an in-progress development program. Key developments required and recommendations for continued development are presented. Comments regarding projected plant performance are presented.

The products, having nil sulfur, nitrogen, and particulate matter, represent premium grade fuels from an environmental standpoint. They also have characteristics which make them attractive as potential feedstocks for high value petrochemical and chemical manufacture.

Details of the design, operating efficiencies, and economic projections are presented in this report.

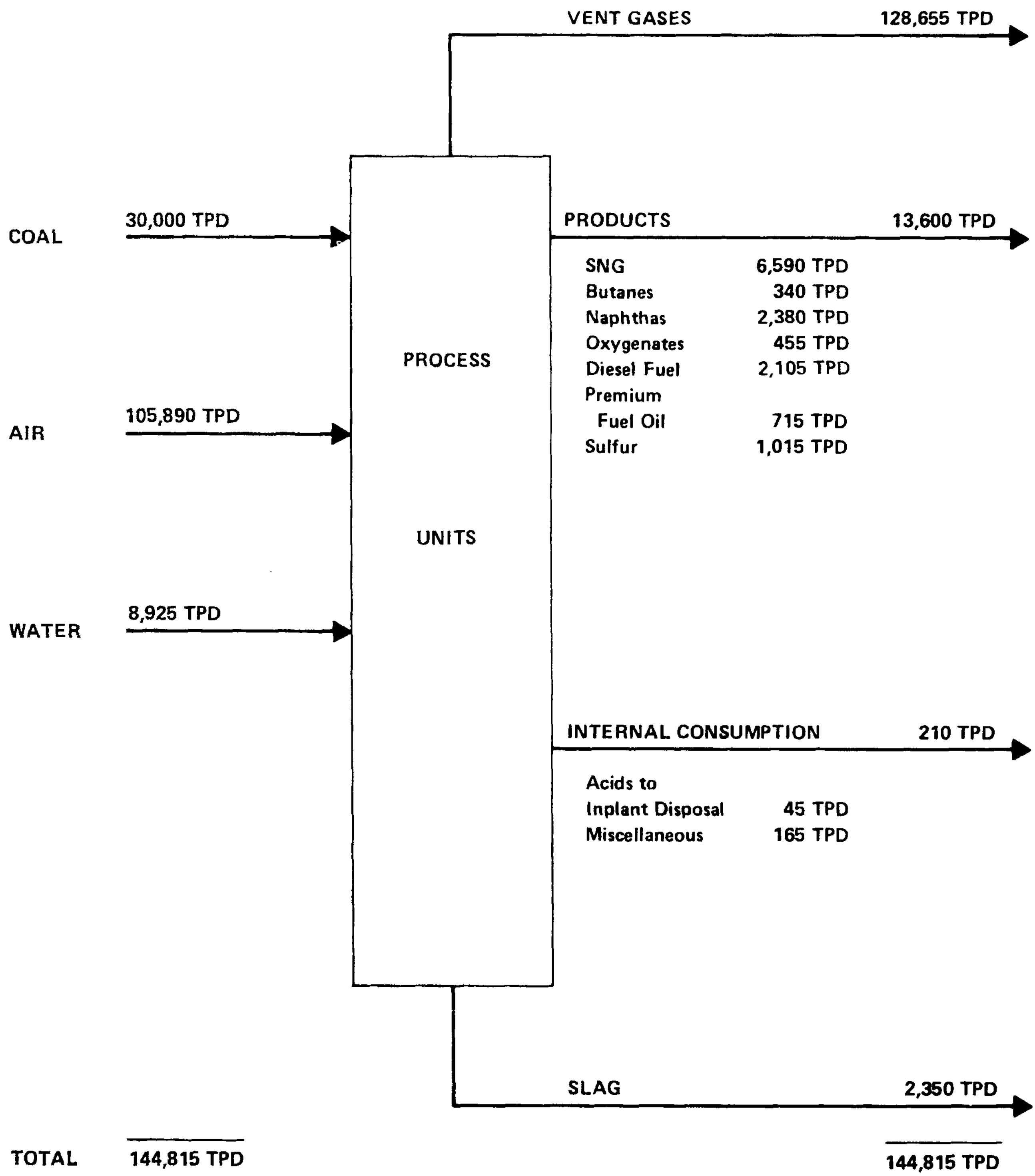

Figure 8-1 - Overall Material Balance
Process Units

COLLECTED WORK NO. 17

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COMPARATIVE ECONOMICS OF COAL CONVERSION PROCESSES

GENERAL PAPERS

CONVERSION OF COAL TO LIQUIDS BY FISCHER-TROPSCH AND OIL/GAS TECHNOLOGIES

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INTRODUCTION

Conversion of coal to liquid and gaseous fuels as well as chemical products has been practiced on a commercial scale in several areas of the world. Projections of U.S. supply and demand balances for crude oil and natural gas to the year 2,000 indicate that coal conversion plants are a candidate in the U.S. for production of environmentally acceptable liquid and gaseous fuels. To be competitive with alternative energy sources, second generation production complexes for coal conversion should be large, efficient, simple and reliable.

This paper describes the characteristics and projected economics for two candidate second-generation technologies, "Oil/Gas" and a U. S. version of Fischer-Tropsch.

The term "Oil/Gas" was coined during the 1974 Project Independence Blueprint campaign. The process uses a type of coal hydroliquefaction similar to SRC II, with reaction severity designed to produce a significant amount of light hydrocarbons. These are in turn processed to yield substitute natural gas (SNG) as a prime product. Liquid products include LPG, naphtha, and fuel oil.

The suggested U.S. version of Fischer-Tropsch incorporates flame-sprayed catalyst on extended heat-exchanger surfaces yielding several potential advantages including increased thermal efficiency. Flame-sprayed catalyst systems have been under development by what is now the Pittsburgh Energy Research Center (PERC) for about 15 years.

The information presented here is based primarily on conceptual designs and economic evaluations prepared by The Ralph M. Parsons Company for the Major Facilities Project Management Division of Energy Research and Development Administration - Fossil Energy (ERDA-FE).[1,2] The conceptual design given for each process incorporates certain process and equipment items now under development, primarily within ERDA programs. The designs are intended to define the potentials for second generation coal conversion complexes incorporating results of in-progress development work. In concept, these complexes might be constructed and operated in the mid-'80's to mid-'90's.

This paper will describe the processing, projected product characteristics, and projected economics for the Fischer-Tropsch and Oil/Gas complexes. These factors will then be compared, recognizing that each produces significantly different products. In addition, the Oil/Gas design[1] will be extended by hydrotreating fuel oil to produce lower percent sulfur products at increased cost, to further illustrate the flexibility of the technology.

Each of these conceptual designs represents only one of numerous possible configurations. For a given industrial application with a defined coal source and required product mix, the design would in actual practice be tailor-made for that particular case.

OIL/GAS

Design Criteria

Preliminary design criteria have been published.[3] Key elements of the completed conceptual design are:

Plant Location	Eastern region of the U.S. Interior Coal Province, which includes portions of the states of Illinois, Indiana, and Kentucky.
Coal Source	Illinois No. 6 seam coal produced in a captive surface coal mine.
Capacity	Approximately 47,000 tons per day (TPD) of run-of-mine (ROM) coal which is cleaned, washed and sized to produce about 36,000 TPD of coal feed to the process plants. All daily figures are in stream days. Products include about 165 million standard cubic feet per day (MM SCFD) of SNG and approximately 75,000 barrels per day (BPD) of liquids consisting of LPG, naphtha, and fuel oil.
Plant Availability	The plant is considered to operate at capacity 330 stream days per year, resulting in an availability factor of 90.4%.
Characteristics	The complex is a grass roots facility which captively produces all utilities and oxygen requirements. All effluent streams are treated to meet environmental standards.
Raw Material and Product Storage	Facilities are provided for a 14-day coal inventory and a 30-day liquid product inventory.

Facility Description

An artist's conceptual drawing is presented in Figure 1. A photograph of a model of the complex is shown in Figure 2. The complex would occupy approximately 600 acres, exclusive of the coal mine. Plant population is about 2,350 people. About 17,500 gallons of water per minute would be drawn from the source river.

Coal Mine	The mine is an integrated strip mine with five separate areas or mining units to produce 47,000 TPD of ROM coal operating 350 days per year. The average overburden is 60 feet and average coal seam thickness is 5 feet. The primary overburden removal is with 170 cubic yard draglines. The ROM coal would pass through a primary separation step located in the mining area

and then be transferred by conveyor to a coal preparation plant area where it is cleaned and sized to produce feed coal to the process plant.

Over the 20-year project operating life, approximately 57 square miles would be mined out.

Process Plant

A process block flow diagram is shown in Figure 3.

Key to the process is the SRC II hydroliquefaction step. Here, 20,000 TPD of cleaned, sized feed coal is slurried in coal-derived recycle solvent; two-thirds of the solvent is unfiltered and contains undissolved coal and ash, while the remaining one-third has been filtered to remove the solids. The coal slurry is pumped to 2,050 psig, mixed with hydrogen, preheated to 700°F, and reacted in the dissolver vessel. The dissolver product passes through a pressure let-down system with the resulting liquid phase going to a low pressure fractionator. Fractionation products are naphtha, light distillate used as fuel oil constitutent, heavy distillate used as filer wash oil and as a product fuel oil constitutent, and the bottoms which contain solids. The bottoms are split; about half are recycled to the feed coal slurry system and the remainder goes to the filters.

The naphtha is hydrotreated to produce saleable product. The light distillate, a portion of the heavy distillate, and the filtrate are combined to form the product fuel oil.

Gases emitting from the dissolver pressure let-down system, fractionation, and the naphtha hydrogenation steps are combined and fed to a monoethanolamine (MEA) acid gas removal system to take out the hydrogen sulfide, carbon dioxide and carbonyl sulfide. The resulting sweet product gas is then processed in a cryogenic unit for hydrocarbon recovery/separation as described below. Sour acid gas is sent to a sulfur plant which removes the sulfur-containing contaminants and produces saleable sulfur.

In the SNG and LPG production train, sweet gas produced in the MEA system is dried with molecular sieves and then sent to a cryogenic unit. Here 98.5 volume percent hydrogen is recovered. A portion of this hydrogen stream is used to methanate residual carbon monoxide. Then the high purity hydrogen is fed to the naphtha hydrotreater while the remainder of the hydrogen stream is recycled to the coal dissolving step. Methane-rich gas produced in the cryogenic unit is compressed, cooled to remove condensible fractions, and then passed through a zinc oxide guard chamber to reduce the hydrogen sulfide content. It is then processed in a final methanation unit and sent to the SNG product line. Ethane and heavier fractions produced in the cryogenic unit are fractionated to remove ethane and some propane overhead which is mixed with final methanator product to produce specification grade SNG, which is compressed to 1,000 psig for delivery. Remaining propane and heavier material is separated into propane LPG as an overhead product and a bottoms product. Bottoms are debutanized to produce butane LPG as an overhead product, leaving pentane-and-heavier bottoms which are fed to the naphtha hydrogenation unit. Butane LPG is hydrotreated and transferred to product storage.

The make-up hydrogen stream for the coal dissolving step is produced in a coal-fed gasifier operated at about 1,000 psig. Significant methane is produced at this pressure. The gasifier is a two-stage entrained slagging type. Solids are removed from the gasifier effluent gas stream and the hydrogen-to-carbon monoxide ratio adjusted in a sour shift conversion unit. The shifted gas is processed in a physical solvent acid gas removal system to produce a sweet gas for feed to the dissolver section, a hydrogen sulfide-rich gas stream for feed to the sulfur plant, and a carbon dioxide-rich vent gas stream. The Rectisol process was used as a representative process.

A fuel gas gasifier system is included to generate the necessary steam and power to operate the complex. This gasifier is fed by the dried dissolver filter cake plus coal. The filter cake is previously dried to recover the wash solvent as a saleable product. Fuel gas generated in the gasifier is treated in an acid gas removal system to remove hydrogen sulfide and carbon dioxide before passing to the power and steam generation section.

Power and Steam Generation

The in-plant produced fuel gas is used to produce electrical power in two condensing turbine generator units with three extraction points. Four steam boilers are also included. The utility system is closely integrated with the process plant operation.

Material Balance

The overall material balance for the process plant is shown in Figure 4. Material inputs consist of coal, water, and oxygen (from the air separation plant). The coal amounts to about 36,000 TPD. Saleable products, including fuels, sulfur and ammonia, add to approximately 19,000 TPD.

Energy Balance

The energy balance is depicted in Figure 5. The projected thermal efficiency, coal to saleable products, is about 77%.

FISCHER-TROPSCH

Design Criteria

Preliminary design criteria have been described.[4] Key elements of the completed conceptual design are:

Plant Location — Eastern Region of the U.S. Interior Coal Province.

Coal Source — Illinois No. 6 seam coal produced in a captive surface coal mine.

Capacity — Approximately 40,000 TPD of ROM coal will be mined and 30,000 TPD of cleaned, sized coal will be fed to the process plant. The products will have an energy value of approximately 525 billion Btu per day consisting of 260 MMSCFD of SNG and approximately 50,000 BPD of liquid products which are LPG's, light and heavy naphthas, diesel fuel, fuel oil and oxygenates.

Plant Availability	350 stream days per year; availability factor = 90.4%.
Characteristics	Grass roots facility producing all utilities plus oxygen and treating all effluent streams to meet environmental standards.
Raw Material and Products Storage	Fourteen-day coal storage and 30day liquid product storage.

FAcility Description

The complex is depicted in the artist's conceptual drawing shown in Figure 6 and a photograph of model of the complex is presented in Figure 7. Land area required for the complex, without the coal mine, is about 500 acres. Plant population is about 2,100 people. Approximately 12,000 gallons per minute (GPM) of water would be required.

Coal Mine

As in the Oil/Gas design, a strip mine with an average overburden of 60 feet and average seam thickness of 5 feet would produce the required 40,000 TPD of ROM coal. The mine would consist of four integrated mining faces. The primary separation and coal preparation units are similar to those previously described for the Oil/Gas complex with the exception that the ground coal has a smaller particle site; minus 20 mesh by 0 for Fischer-Tropsch vis-a-vis 5% plus 20 mesh, 25% minus 200 mesh for the Oil/Gas plant.

Process Plant

All of the feed coal is fed to two entrained slagging-type steam oxygen gasifiers operated at approximately 470 psig. Gasifier effluent gas stream is exhaustively cleaned to remove solid particles. The ratio of hydrogen to carbon monoxide in the cleaned gas is increased by subjecting about 50% of the gas stream to a shift conversion reaction; the H_2/CO ratio is thereby adjusted to the target value of 1.45. Shifted gas is then fed to an acid gas removal unit where it is contacted with a physical solvent to remove the hydrogen sulfide, carbon dioxide and organic sulfur compounds. The Selexol process was used as a representative process for this design. The absorbed acid gases are stripped for further processing; the hydrogen sulfide is converted to saleable sulfur in the sulfur plant and the CO_2 stream is vented. Sulfur content of the cleaned syngas is reduced to about 0.1 part per million, volume (ppmv).

Cleaned syngas is fed to the Fischer-Tropsch synthesis unit at about 400 psig and 570°F. It first passes through zinc oxide guard chambers to remove trace quantities of sulfur compounds. Then it is processed in 18 parallel synthesis reactors designed for isothermal operation. The reactors have flame-sprayed iron catalyst deposited on the external surface of extended surface heat exchangers. Reaction takes place on the shell side and 1,250 psig steam is generated on the tube side by the heat of reaction. Shift and methanation reactors have a similar geometrical design but differ in the composition of the flame-sprayed catalyst.

Fischer-Tropsch reactor feeds contain a ratio of recycle to fresh feed of approximately 1.4. Extensive heat exchange is used to maintain a high plant thermal efficiency.

Fischer-Tropsch products go to a liquid product recovery unit to recover light hydrocarbons from the Fischer-Tropsch gas and to fractionate the liquids into the product streams.

Two gas streams are recovered and fed to the methanation unit which produces SNG. One consists of a mixture of residual lean gas after absorption of the C_3^+S in a presaturated lean oil stream and a CO-rich stripper overhead product produced by stripping a lean oil fractionator overhead stream. This mixed stream is fed to the first methanation stage. It contains gases produced in the Fischer-Tropsch reactor, including methane and some C_2's and C_3's to increase the heating value of the SNG. An additional feed stream, which goes to the second-stage methanator, consists of C_3's and C_4's which are produced in a depropanizer in the liquid product refining train; they serve to increase the heating value of the SNG.

Fischer-Tropsch liquids are preheated and fed to a lean oil fractionator where light ends are removed overhead for further processing and feed to the methanator section as described previously. The bottoms are fed to the fuels vacuum fractionator where the heavy naphtha, diesel oil and heavy fuel oil are produced. Naphtha is removed as an overhead product. Diesel oil is withdrawn as a side stream and is steam-stripped to obtain the flash point specification. Heavy fuel oil is produced by steam stripping in the bottom section of the fractionator, cooled, and sent to storage.

Light naptha is produced in a naphtha stabilizer fed by the bottoms from the depropanizer. C_4 LPG's are recovered as overhead from the stabilizer.

Oxygenate produced in the Fischer-Tropsch reactor, containing a high alcohol content, are recovered and refined. Feed to the oxygenate recovery system is produced in a water extraction of the Fischer-Tropsch liquids. This feed is preheated and the oxygenates taken overhead from a fractionator with the bottoms returned to the extraction system. A hot alcohol-salt solution, produced by caustic neutralization of the Fischer-Tropsch reactor effluent to destroy acids produced in the reaction, is stripped and the oxygenates recovered as an overhead product are also fed to the oxygenate fractionator previously discussed. The stripper bottoms are evaported to produce a concentrated salt solution for disposal and a consensate stream used as boiler feed water.

Product SNG is produced in the methanation section. The primary feed is sulfur-free stripped gas produced in the liquid product recovery section. The methanation section consists of a first-stage recycle reaction unit containing three methanators in parallel, and a second-stage one-pass finishing reactor.

Feed gas to the first stage methantor is mixed with 1.25 parts of recycle gas, preheated to about 570°F, and reacted at 380 psig in isothermal reactors of design similar to those used for the Fischer-Tropsch reaction. A flame-sprayed nickel catalyst is deposited on the outside surface of a finned tube heat exchanger and the high heat of reaction removed by boiling dowthern in the tubes -- the hot dowthern in turn is used to generate 1,300 psig steam for use in the plant utility system. Reaction conditions in the first-stage methanator favor CO methanation to assure that the product SNG does not contain more than 0.1 mol% CO. Product from this first-stage methanator is cooled, condensate removed, and about three-fourths of the gas recycled with the remainder going to the second-stage methanator which is an adiabatic fixed-bed radial-flow reactor using a pelleted, reduced, nickle-type catalyst. Here the CO_2 is methanated; it will also methante CO if a breakthrough should occur in the first stage. The CO_2 content of the product SNG is maintained below 2.5%.

The product from the second-stage methanator has a higher heating value of about 910 Btu/SCF. This is combined with the vaporized mixed light hydrocarbon stream produced in the liquid product recovery section and fed to a hydrotreater for saturation of alkenes by the residual hydrogen in the stream. The product SNG stream is cooled, condensate removed, compressed, dried, and fed to the product pipeline at 1,000 psig.

POWER AND STEAM GENERATION

The process produces all steam required for operations, heating, and power generation. Therefore, conventional steam boilers are not provided for normal operation. A start-up boiler is provided.

Electrical power is generated by four 120-megawatt extraction steam turbine generators. These generators provide all power required for operation of the complex plus approximately 140 MW for sale.

MATERIAL BALANCE

Overall material balance for the process units is presented in Figure 8. Results indicate that approximately 13,000 tons per day of saleable fuel products plus sulfur are produced from 30,000 tons per day of cleaned, sized feed coal.

ENERGY BALANCE

Energy balance is summarized in Figure 9. Estimated thermal efficiency in converting coal to saleable products is approximately 70%.

PRODUCT CHARACTERISTICS

Projected product characteristics for the Fischer-Tropsch and Oil/Gas conceptual designs are summarized in Table 1. These have been projected based on review and analysis of product characteristics reported by process investigators for similar, but not identical, process conditions[5,6] plus minor adjustments to reported product characteristics using the characteristization factor to assure consistency with the basic data. For more radical adjustments to reported product characteristics as a result of subsequent treatment, for example, hydrogenation, reference was made to published work[8,9] in this area to establish change of characteristics resulting from treatment. There are not yet reports of production-analysis-functional product testing of large quantities of the naphtha, diesel fuel, and fuel oil streams. However, the projection of these characteristics based on analysis of existing data and comparison of expected values based on analogy to other coal-derived liquids and similar crude oil-based products provides a basis for projecting comparative results for these two technologies and defining incentives for pilot plant production to permit confirmation or modification of the projections.

The most significant differences are that the Fischer-Tropsch liquid products contain nil sulfur, nitrogen and particulate matter, and are composed primarily of aliphatic compounds, while the Oil/Gas products contain sulfur, nitrogen, and solids and consist primarily of aromatics. The Fischer-Tropsch liquids therefore have higher potential for use as petrochemical feedstocks and for fuel applications with stringent environmental restrictions. Oil/Gas products show promise for us in gasoline manufacture and for selected fuel applications. Additional comments will be presented later regarding possible market values of these products.

FIXED CAPITAL INVESTMENTS

All economics are expressed in Fourth Quarter 1976 dollars.

The projected fixed capital investments (FCI) for the two conceptual complexes are compared in Table II. The results indicate that the Fischer-Tropsch complex would require a FCI of approximately 1.55 billion dollars to produce about 85,000 barrels of fuel oil equivalent per day (BOE/D); the FCI per BOE/D is therefore about $18,000. The Oil/Gas complex would require a FCI of about $1.3 billion to produce approximately 110,000 BOE/D for a FCI per BOE/D of about $12,000.

A comparison of the relative costs of the separate sections of the complex is shown in Table III. A significant contributor to the higher FCI per BOE/D for the Fischer-Tropsch plant lies in the gasification section where the cost of the oxygen plants and gas cleanup are much higher. Note that the FCI's for the conversion sections, per daily barrel of oil equivalent, for the two complexes are about equal.

TOTAL CAPITAL INVESTMENTS

Total capital investments are presented in Table IV. Total capital includes fixed capital investment, initial catalyst and chemicals, start-up costs, construction financing, working capital, and land/rights of way. Projected total capital requirements are 2.0 and 1.7 billion dollars for the Fischer-Tropsch and Oil/Gas complexes, respectively. Example construction financing costs are presented in each case.

Estimated time to mechanical completion was approximately 57 months in each case. This included design, engineering, procurement and construction.

OPERATING COSTS

Projected annual operating costs for the complexes are given in Table V. The operating costs include royalty allowance of $1.50 per ton of cleaned, sized coal produced.

Projected annual operating costs are 205 million dollars for both the Fischer-Tropsch and Oil/Gas complexes, respectively. For analytical purposes, the complexes were divided into cost centers.

REQUIRED PRODUCT SELLING PRICE

Average required product selling price was projected for three project financial structures. In all cases, the project operating life was 20 years.

- 100% equity capital
- Borrowed capital: 65% of the total investment borrowed at 9% interest, with the principal repaid in equal installments over a 20-year project operating term; all working capital borrowed at 9% interest for the 20-year term; a loan commitment fee of 0.75% on funds not drawn down during the construction period.
- A nonprofit (0% discounted cash flow rate of return) or break-even boundary case.

A 12% discounted cash flow rate of return (DCF) was selected as a base case, and the revenue required to achieve this DCF calculated for each financial structure. Required average product selling price was then calculated using the required revenue and the quantity of energy products produced.

Results are summarized in Table VI. Here we see that for the 65/35 debt/equity financial structure, the projected average required product selling prices, fourth quarter 1976 basis, (RPSP) are $2.55 and $1.95 per million Btu's for the Fischer-Tropsch and Oil/Gas cases, respectively. The 100% equity financing cases are about 30 percent higher in each case. The breakeven cases are about $1.50 and $1.20 per million Btu's, respectively.

In dollars per barrel, the 65/35 debt/equity case RPSP's would be about $15.25 and $12.00; this is based on an arbitrary 6 million Btu per barrel reference value. A key factor in the economic projections is the inclusion of large captive coal mines in the complexes.

SENSITIVITIES

Sensitivities of the average required product selling price to changes in key economic parameters are shown in Table VII. The RPSP is most sensitive to changes in fixed capital investment. To illustrate, for Fischer-Tropsch a 10% reduction in fixed capital investment would result in an 8.7% reduction in RPSP for the 100% equity case. The sensitivities to operating costs are in the range of 15-20%.

Effect of variations in DCF on the RPSP is presented in Figure 10 for the 65% debt case. Sensitivity is greater for the 100% equity case, which is not shown.

POSSIBLE PRODUCT MARKET VALUES

A brief assessment of possible product market values and the effect of the resulting project revenues on profitability was completed. To obtain these possible market values, the project characteristics of the products were compared with those of conventional crude oil-based products. Discussions were held with representatives of fuel producers and consumers and industry reports were reviewed.

Industry representatives strongly qualified their opinions, on possible prices by stating that laboratory and field product performance tests must be conducted before firm dollar values could be assigned to the products.

With the above caveats clearly in mind, possible unit sales values and annual revenues for a fourth quarter 1975 basis are presented in Table VIII for Fischer-Tropsch and Table IX for Oil/Gas; these are taken from the published reports.[1,2] The SNG sales value was based on value allowed for sale of SNG produced commercially from naphtha, and possible values for SNG from coal at that time. These possible sales values are presented to illustrate the effect of product sales value on the economics and also to perhaps stimulate further effort to establish firm product values and marketability.

The Tables VIII and IX possible annual revenues were then updated to a fourth quarter 1976 basis using Federal Energy Administration data which indicated that fuel prices escalated approximately 9 percent from fourth quarter 1975 to fourth quarter 1976.

Results of this second-order exploratory analysis indicate that possible average annual revenues (Fourth Quarter 1976 dollars) are $730 and $560 million dollars for the Fischer-Tropsch and Oil/Gas cases, respectively. Projected DCF's calculated using these revenues and the project structures developed earlier are shown in Table X. To illustrate, for the 65/55 debt/equity case, the projected DCF's for Fischer-Tropsch and Oil/Gas are 27 and 20 percent, respectively. This result indicates the incentive for accurate assessment of the marketability and profitability of synfuel products to be produced in second generation coal conversion plants in the U.S.

ECONOMIC COMPARISON FOR LOW SULFUR CONTENT FUEL PRODUCTS.

Projected sulfur content of the Oil/Gas fuel oil is 0.45. A brief and very preliminary analysis of the effect of further hydrotreating to reduce the sulfur content on cost and product composition was made; this is an extension of the design previously reported.[1] The result provides guidance regarding the costs and implications of producing very low sulfur fuels from coal by Oil/Gas type technology for environmental reasons.

The data basis for predicting process and cost results for hydrotreating the coal-derived liquids is limited. However, some information is available to guide the projections.[8,9,10]

Preliminary process designs were developed for incremental hydrotreating of the Oil/Gas fuel oil. Hydrotreating conditions were nominally 650°F and 2,500 psig with a nickel-molybenum type catalyst. A 6 months catalyst life was assumed for the purpose of this preliminary assessment.

Projected product distribution as a function of fuel oil sulfur content is depicted in Figure 11. With decreasing sulfur content, the amount of fuel oil decreases and the lighter products increase.

Figure 12 presents projections of hydrogen consumptions and Figure 13 shows projected required average product selling price at 12% DCF, 65% debt as a function of fuel oil sulfur content. Also shown on Figure 13 is the projected

RPSP for nil sulfur Fischer-Tropsch products. Results indicate that at about 98% sulfur reduction in Oil/Gas fuel oil, the required product selling prices are approximately equal.

SUMMARY AND CONCLUSIONS

Conceptual designs'economic evaluations for two condidate second generation coal conversion technologies have been completed by the Ralph M. Parsons Company. These are a suggested future version of a Fischer-Tropsch complex, and an Oil/Gas Complex which uses SRC II technology. Each conceptual design incorporated certain process and equipment concepts currently under development. The designs are based on the presumption that these development programs will be successful.

The conceptual complexes process 30,000-36,000 tons per day and produce 85,000-110,000 barrels per stream day of oil equivalent. Projected fixed capital investments for the Fischer-Tropsch (F-T) and Oil/Gas (O/G) complexes are 1.55 and 1.3 billion dollars, respectively; all economics are presented in fourth quarter 1976 dollars. Unit fixed capital investments, expressed as dollars per daily barrel oil equivalent (BOE/D) are about $18,000 and $12,000, respectively.

Projected product characteristics from the complexes differ; Fischer-Tropsch produces primarily aliphatic liquids and Oil/Gas primarily aromatics.

Projected required selling prices to achieve a 12% DCF using a 65% debt, 9% interest case are about $15.25 and $12.00 per equivalent barrel. A second order assessment of possible product sales values has led to the conclusion that DCF's of the order of 20% might be achieved; this is presented to illustrate the incentive to produce and test enough of the synfuels to determine their market values.

Projections of possible costs for hydrotreating a 0.4% sulfur Oil/Gas fuel oil to reduce its sulfur content have been presented. Results indicate that reducing the sulfur content to 0.1% would add an incremental $500 million to the fixed capital investment and reduce the quantity of fuel oil by about 6 percent while increasing the quantities of LPG's and naphtha. A further result is a 15 percent increase in the average required product selling price (RPSP). The average RPSP at this sulfur level is projected to be about 90% of the nil sulfur F-T RPSP. At 98% sulfur reduction in Oil/Gas fuel oil, the RPSP's are about equal. Limited information is available for this hydrotreating step. An incentive exists to develop a firm basis for design and prediction of economics.

Fischer-Tropsch and Oil/Gas coal conversion technologies each offer different advantages and potential problems to be overcome. They must be considered candidates for any future synfuels-from-coal programs.

ACKNOWLEDGEMENT

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10. Ralph M. Parsons proprietary data files.

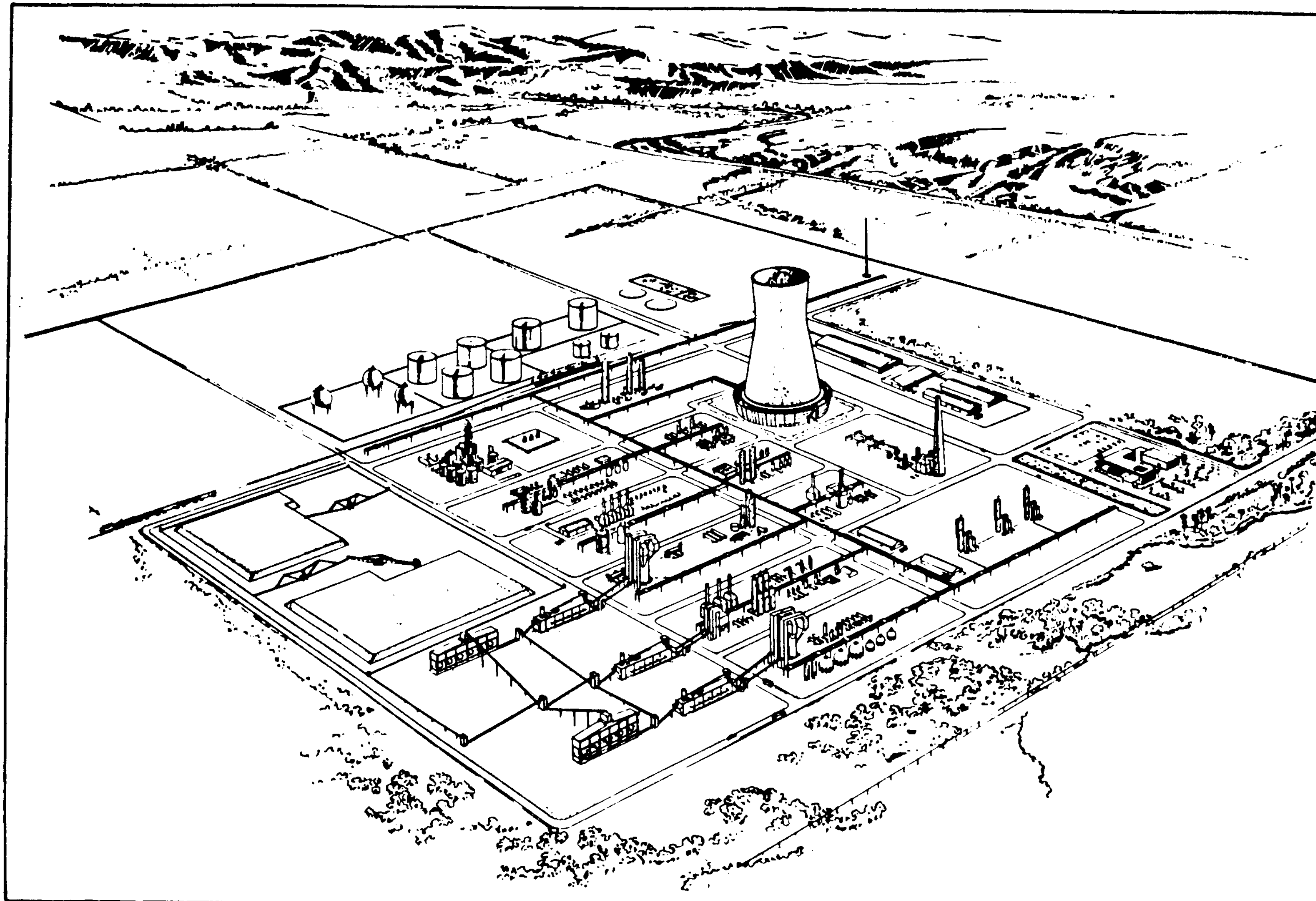


Figure 1 - Artist's Concept, Oil/Gas Plant

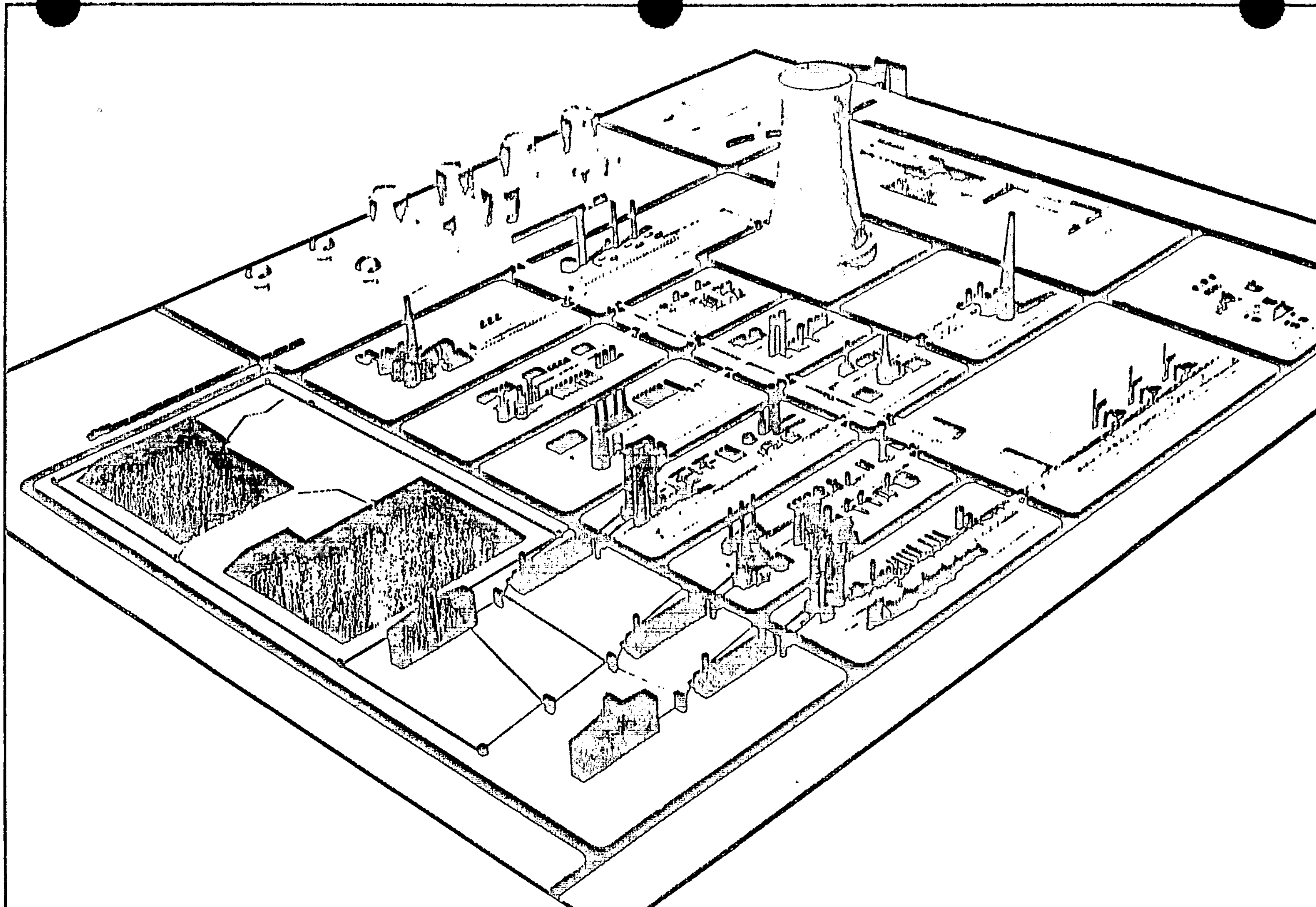


Figure 2 - Model of Conceptual Oil/Gas Plant Design

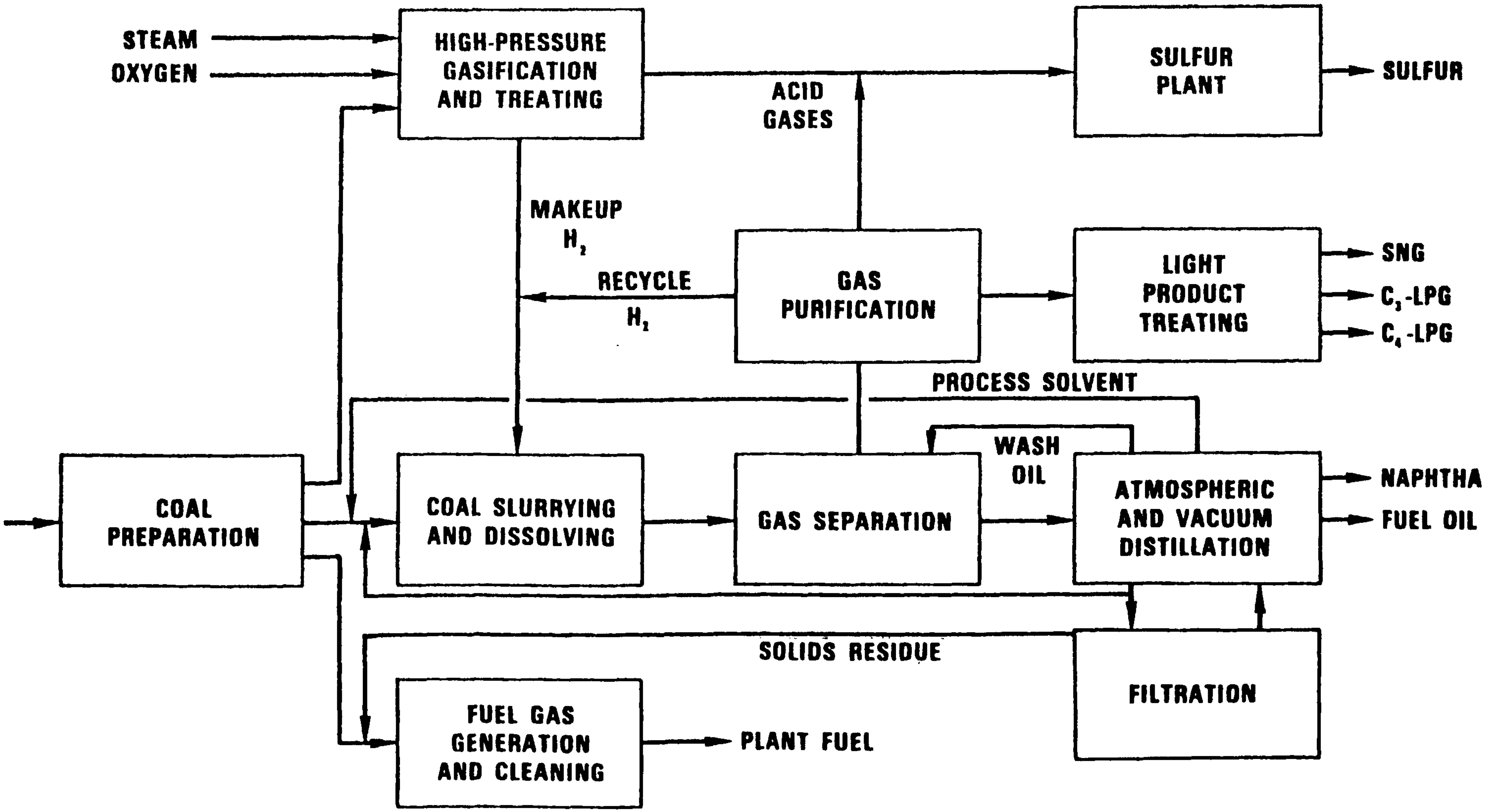
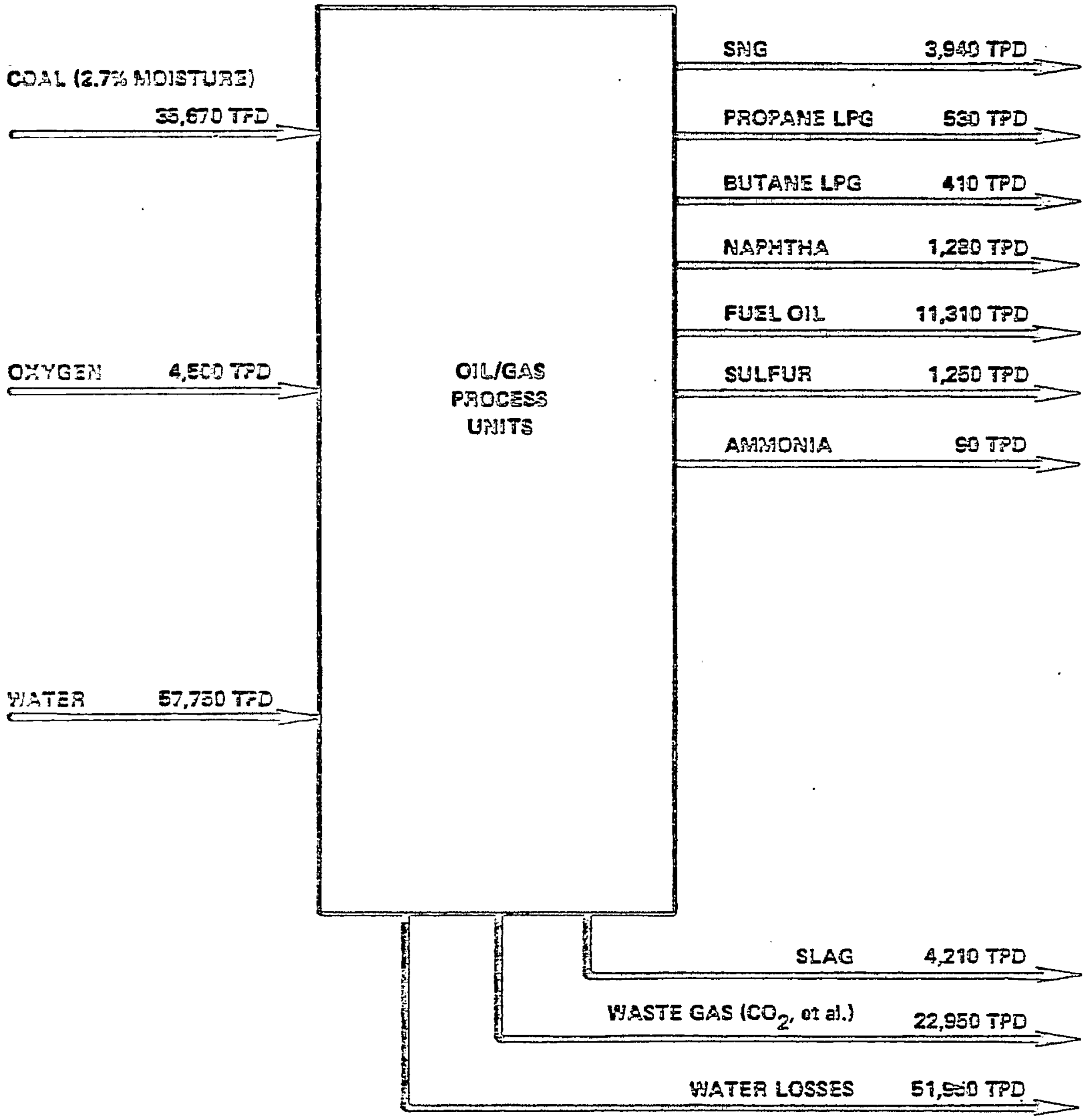


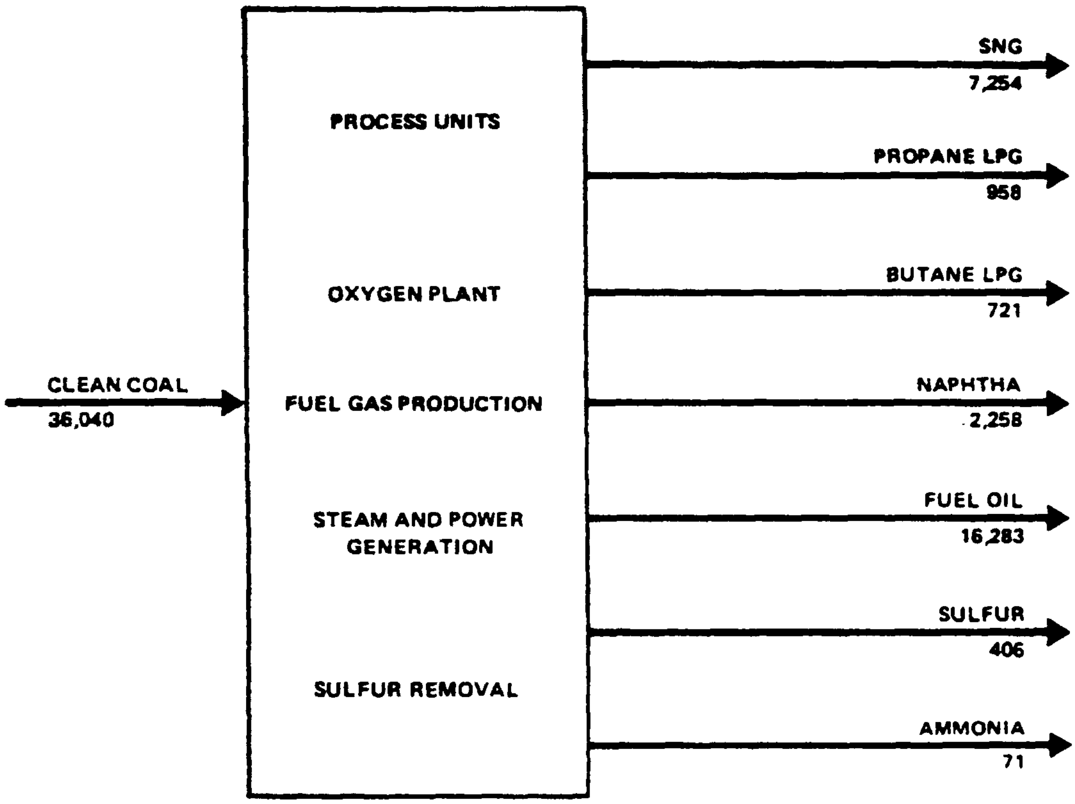
Figure 3 - Oil/Gas Plant
Simplified Block Flow Diagram



TOTAL IN = OUT 97,920 TPD

ALL FIGURES IN SHORT TONS

Figure 4 - Overall Material Balance
Oil/Gas Plant



ALL FIGURES ARE MM BTU/HR, HHV

$$\text{THERMAL EFFICIENCY} = \frac{7{,}254 + 958 + 721 + 2{,}258 + 16{,}283 + 406 + 71}{36{,}040} = 77.6\%$$

Figure 5 - Thermal Efficiency, Oil/Gas Plant

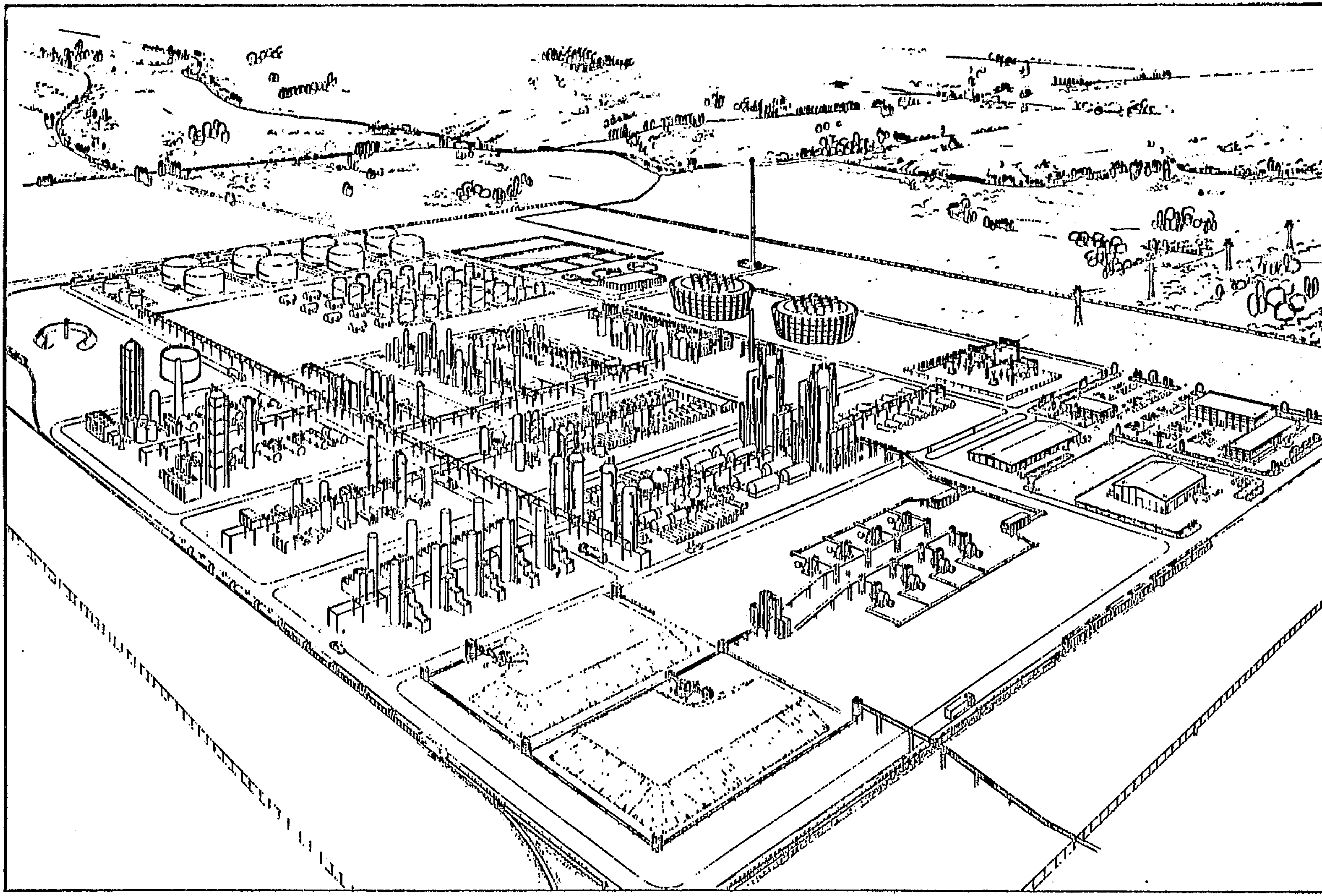


Figure 6 - Artist's Conceptual Drawing
Fischer-Tropsch Plant

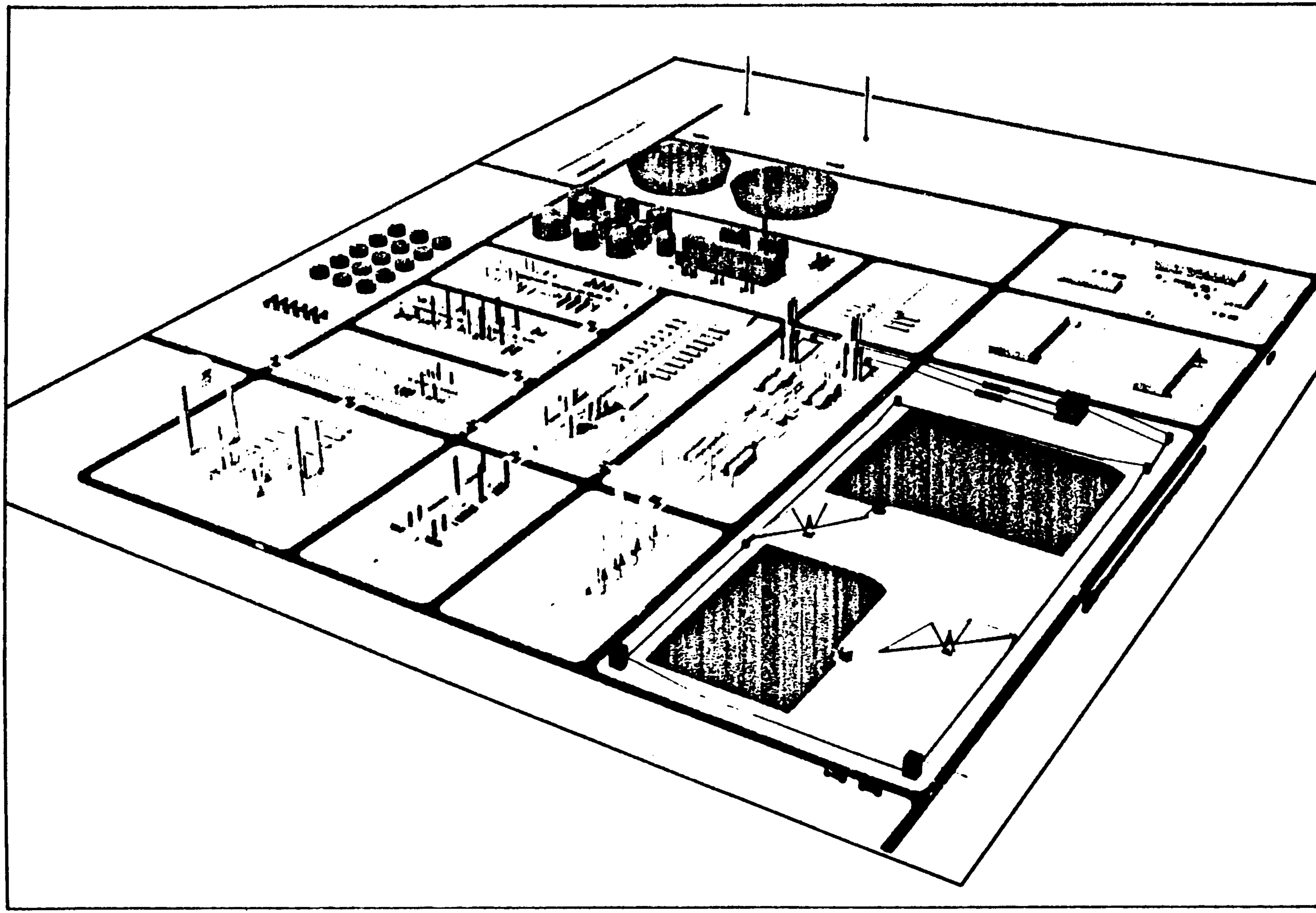


Figure 7 - Model of Conceptual Fischer-Tropsch Plant Design

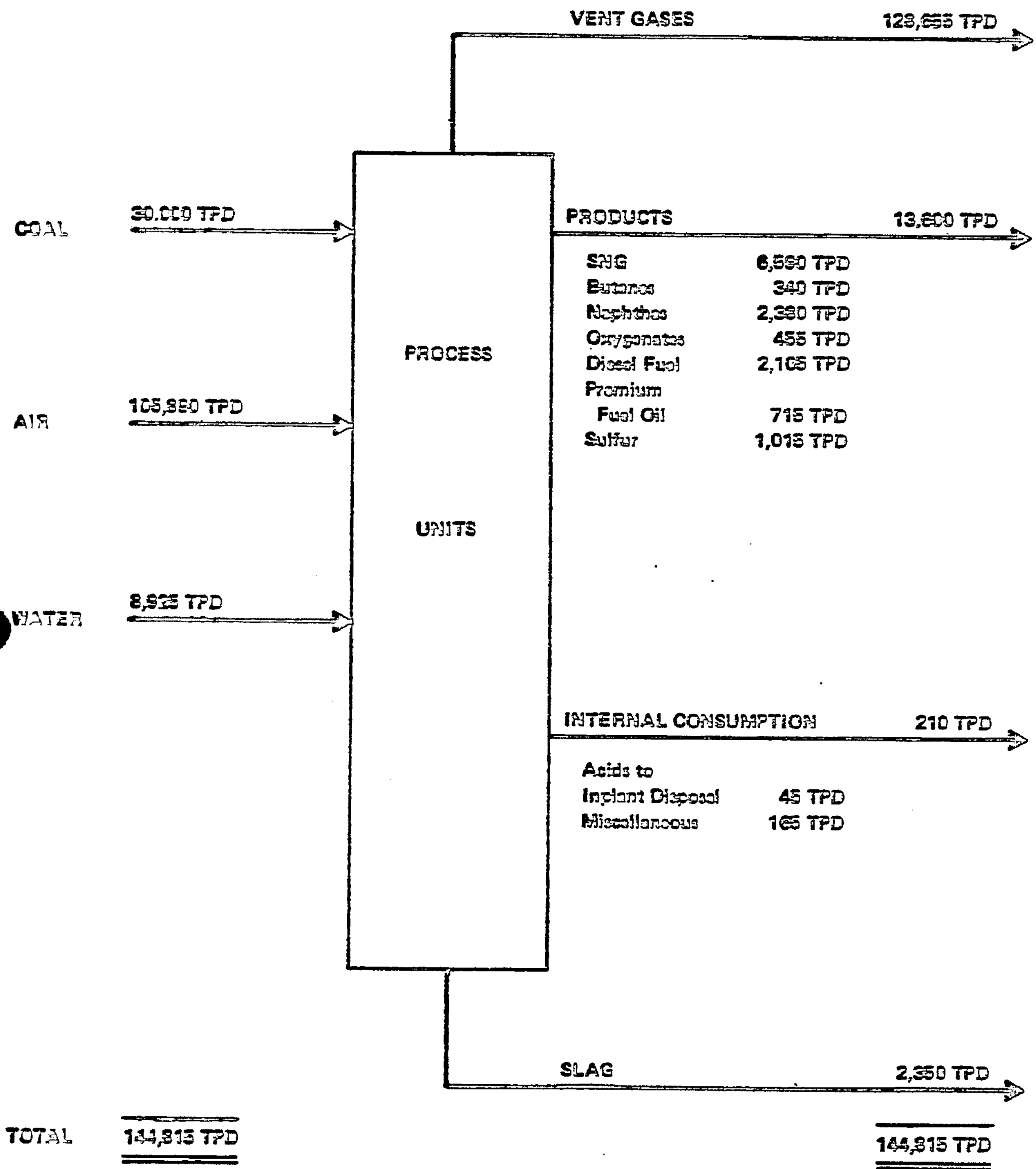
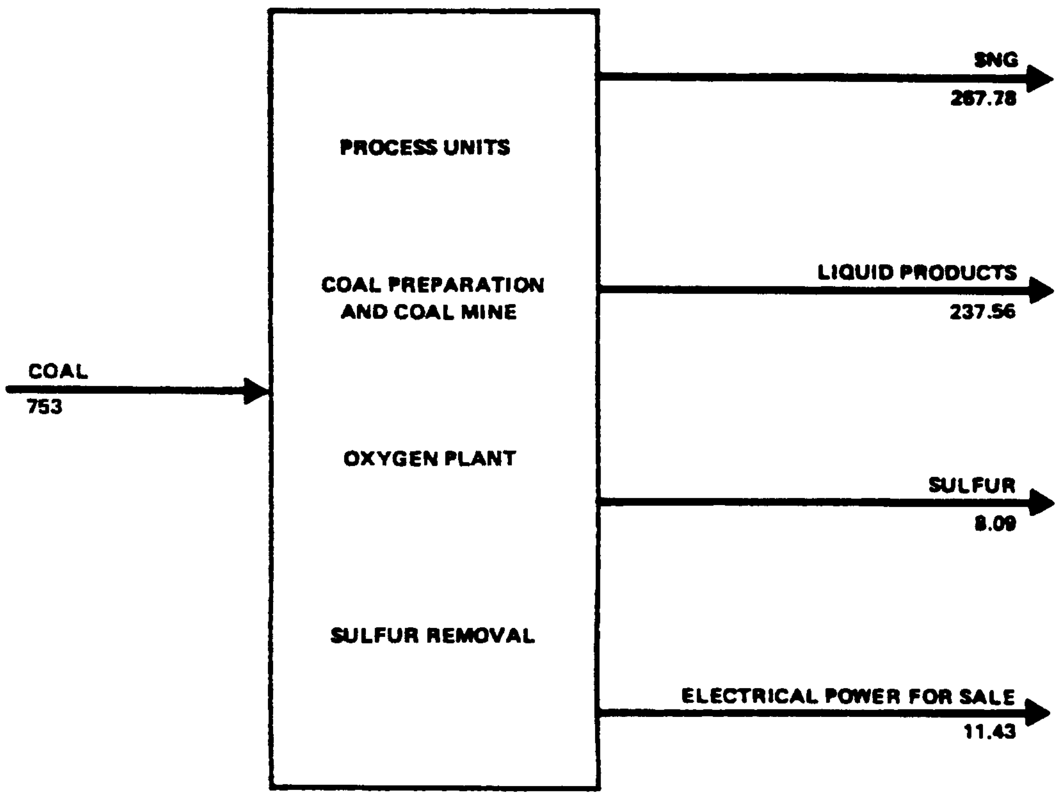


Figure 8 - Overall Material Balance
Fischer-Tropsch Plant



ALL FIGURES ARE MMM BTU/D, HHV

$$\text{THERMAL EFFICIENCY} = \frac{267.78 + 237.56 + 8.09 + 11.43}{753} = 69.7\%$$

Figure 9 - Thermal Efficiency
Fischer-Tropsch Plant

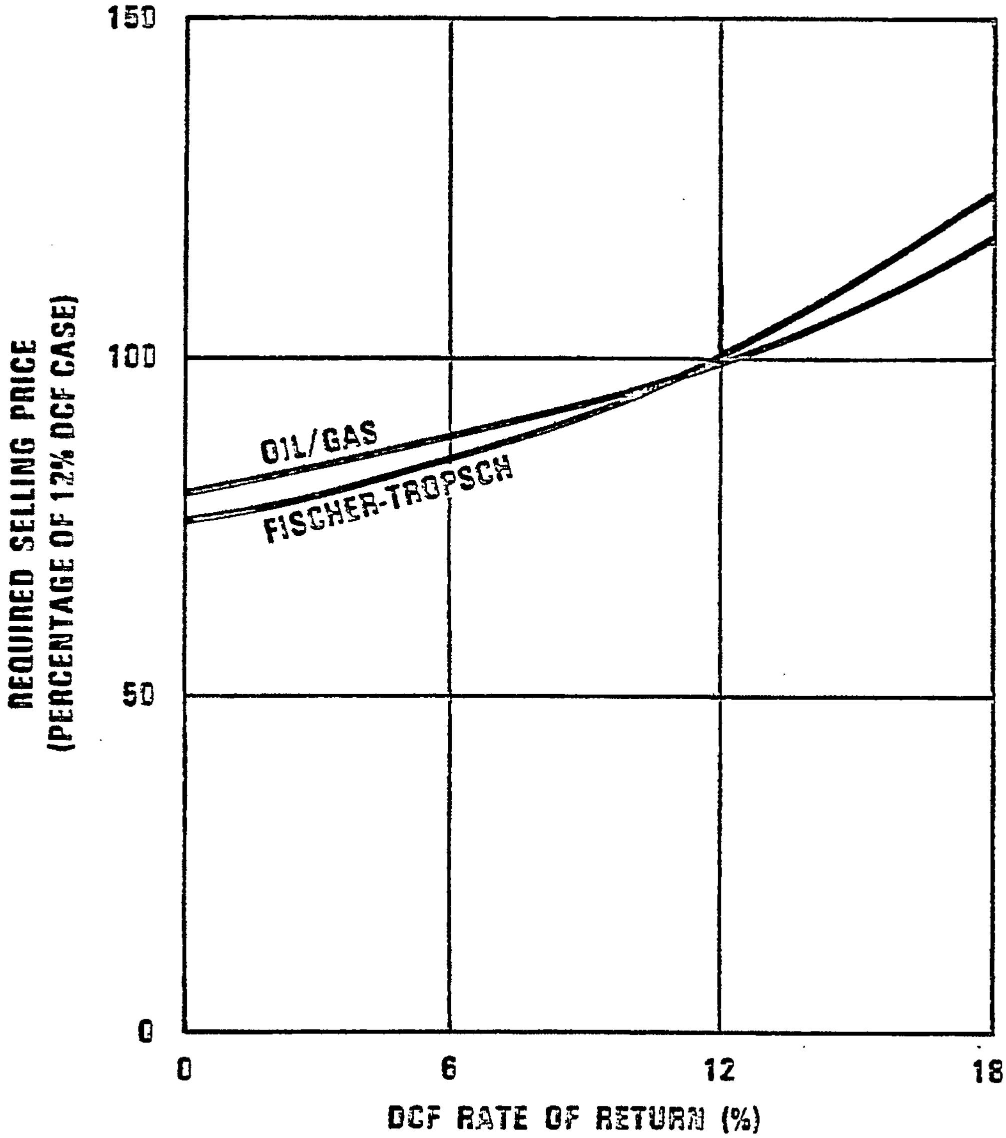


Figure 10 - Sensitivity of Required Product Selling Price to DCF, 65% Debt Case

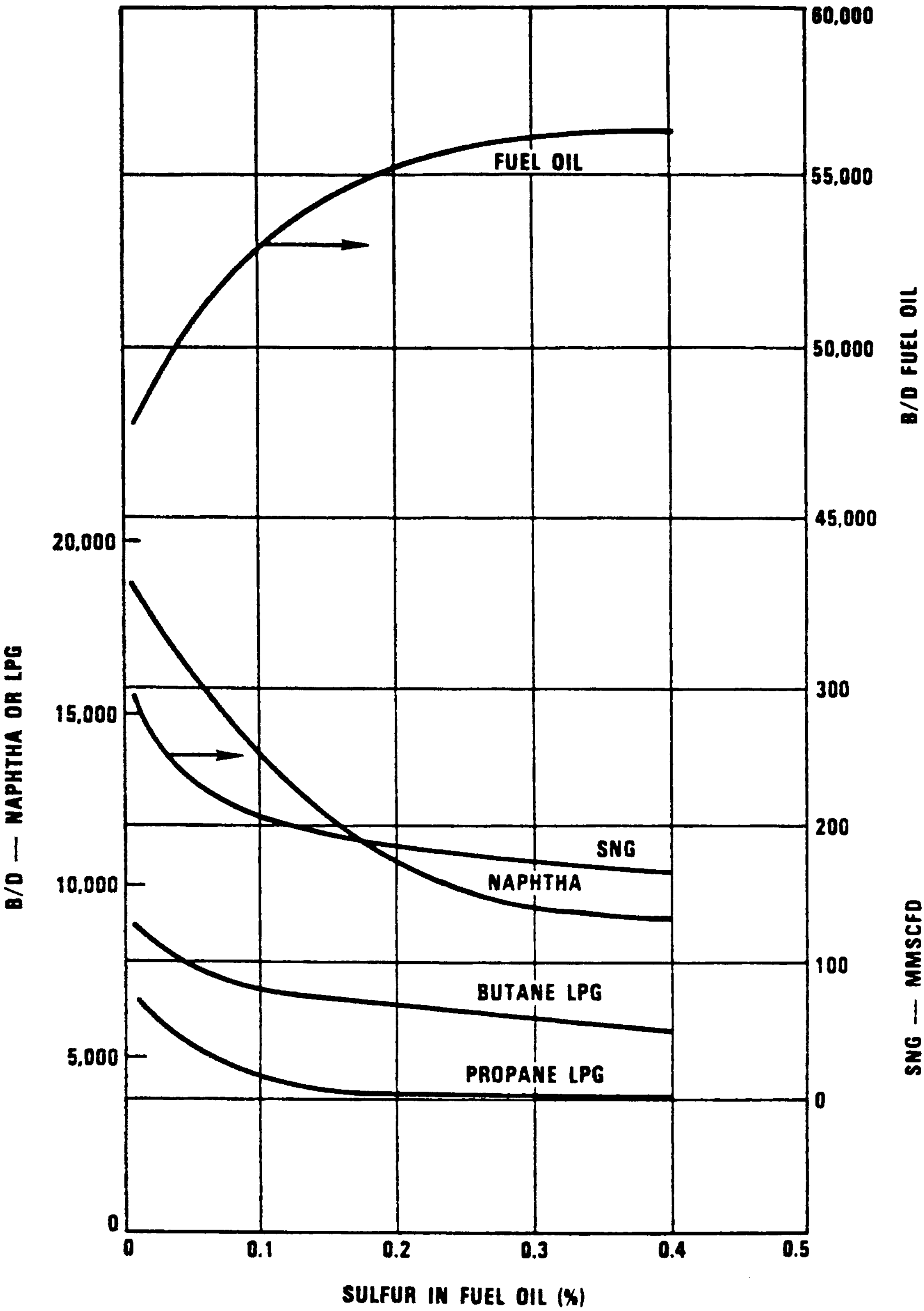


Figure 11 - Projected Product Distribution
Product Yield vs. Fuel Oil Sulfur Content
Oil/Gas Plant

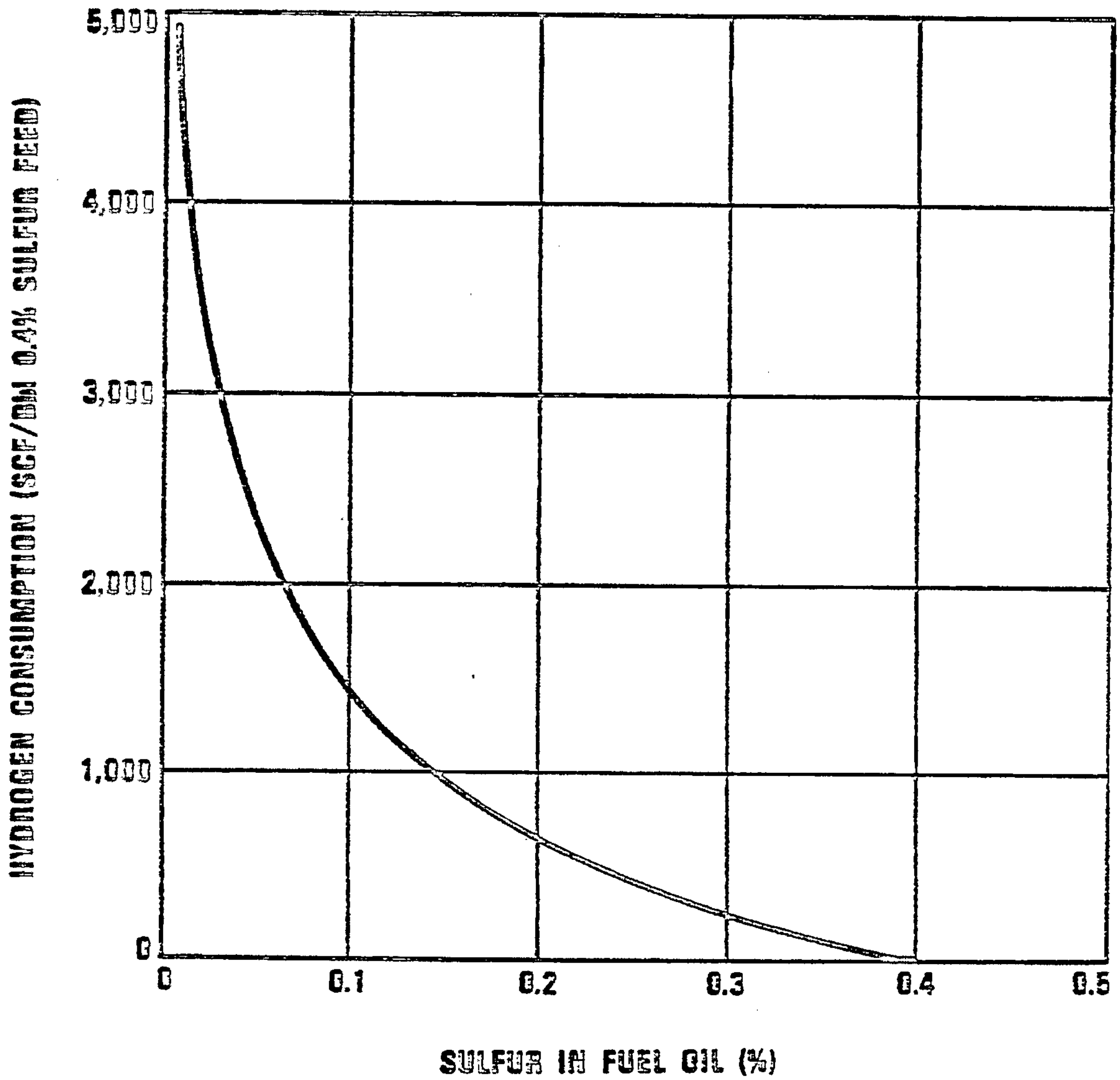


Figure 12 - Projected Hydrogen Consumption
Hydrogen Consumed vs. Fuel Oil Sulfur Content
Oil/Gas Plant

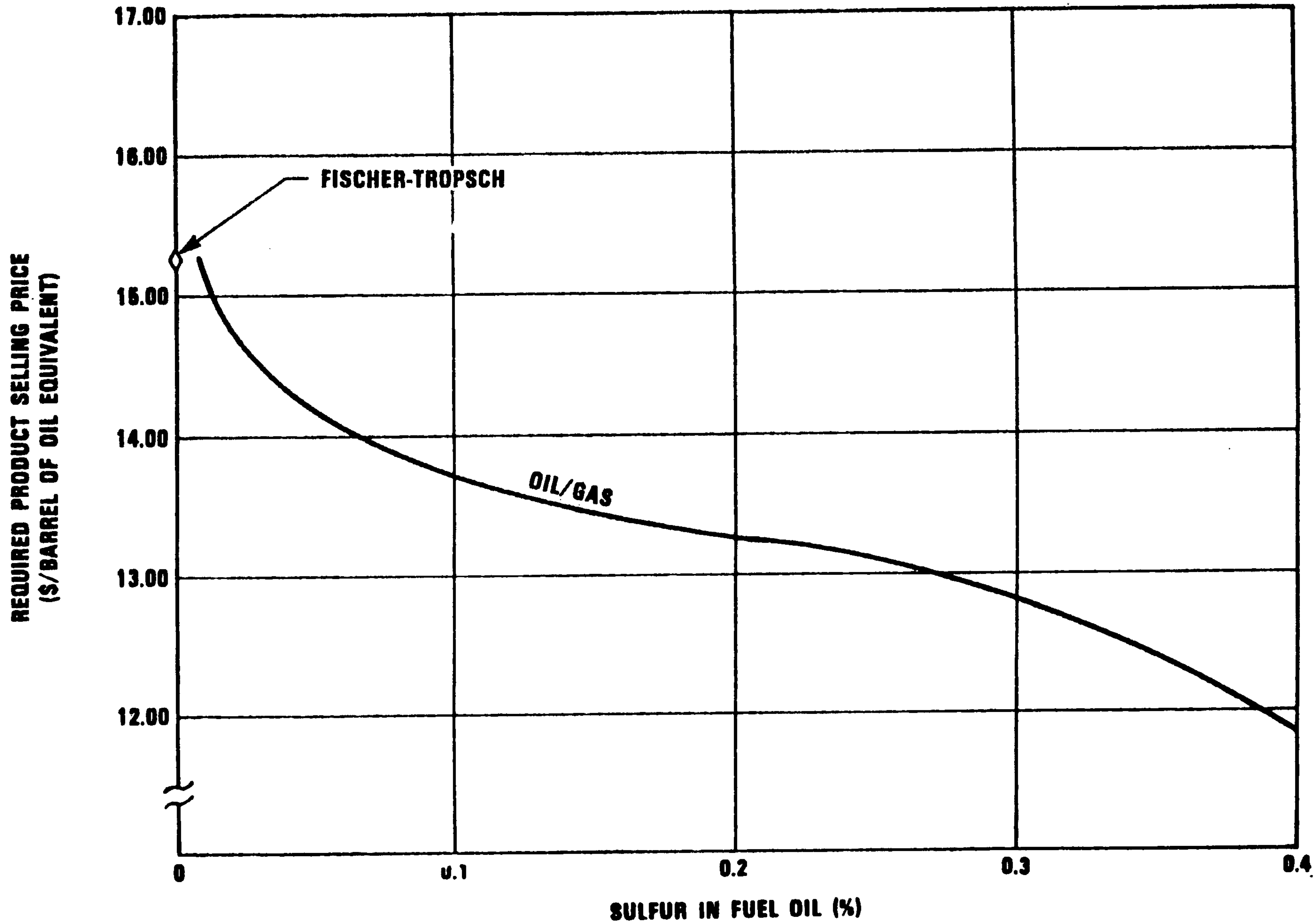


Figure 13 - Required Average Product Selling Price as a Function of the Sulfur Content of Oil/Gas Fuel Oil

Table I. Comparison of Projected Product Characteristics

Product	Projected Characteristics: Fischer-Tropsch	Projected Characteristics: Oil/Gas
SNG	Pipeline Quality	Pipeline Quality
C_3 LPG	--	Propane 210 psia Vapor Pressure
C_4 LPG	Mixed Butane - Butylene 37 psia Vapor Pressure	Mixed Propane-Butane, 70 psia Vapor Pressure
Full Range Naphtha	--	50° API. Gravity C_5 to 380°F ASTM EP High-Naphthene
Light Naphtha	Nil Sulfur 185°F ASTM EP 85.5°API Gravity	--
Heavy Naphtha	Nil Sulfur 300°F ASTM EP 71.3 API Gravity	--
Diesel Fuel	57°API Gravity 60 plus Cetane Number Nil Sulfur, Nil Nitrogen	--
Fuel Oil	41°API Gravity Nil Sulfur Higher Heating Value: 19,900 Btu/lb	-8.2°API Gravity 0.4 wt % Sulfur Higher Heating Value: 17,200 Btu/lb

Table II. Comparison of Fixed Capital Investments (FCI) for Fischer-Tropsch and Oil/Gas Complexes

Barrels Fuel Oil Equivalent/Day (BPOE/D): F-T = 86,000
O/G = 110,000

Description	Fischer-Tropsch $ Millions	Oil/Gas $ Millions
Mine and Coal Preparation		
Mine	175.6	211.6
Coal Preparation	22.0	30.0
Coal Storage	11.2	13.0
Crushing and Drying	13.0	15.1
Subtotal	221.8	269.7
Conversion		
Fischer-Tropsch Synthesis	204.6	--
Oil Recovery and Fractionation	30.5	--
Chemical Recovery	15.9	--
Slurry and Dissolving	--	216.8
Filtration	--	42.0
Distillation	--	31.6
Dissolver Acid Gas Removal	--	20.3
Subtotal	251.0	310.7
Process Gas Production		
Gasification	37.3	45.4
Heat Rec. and Part. Removal	151.2	--
Acid Gas Removal	100.3	47.7
Shift	18.9	59.5
Power Generation	119.6	86.8
Subtotal	427.3	239.4
SNG Separation and Treatment		
Methanation	60.6	0.6
SNG and LPG Treating	--	48.3
Subtotal	60.6	48.9
Product Finishing		
Sulfur Plant	22.1	15.4
Naphtha Hydrogenation	--	9.2
Subtotal	22.1	24.6
Utilities		
Oxygen Plant	305.3	90.2
Instrument and Plant Air	5.6	2.4
Potable and Sanitary Water	0.4	--
Raw Water System	23.8	--
Fuel Gas Gasify	--	71.2
Fuel Gas Acid Gas Removal	--	17.9
Raw Water Treating	--	16.5
Subtotal	335.1	198.2
Environmental and General Facilities		
General Facilities	19.5	37.2
Water Reclaiming	40.4	--
Effluent Water Treating	3.0	5.5
Product Storage	21.2	32.2
Sour Water Stripping	--	5.9
Subtotal	84.1	80.8
Total Constructed Cost	1,420.0	1,172.3
Home Office Costs	140.2	117.2
Sales Tax	28.1	23.5
Total Fixed Capital Investment (FCI)	1,570.3	1,313.0
FCI/(BPOE/D)	18,250	11,950

Table III. Comparison of Relative Fixed Capital Investments of Fischer-Tropsch and Oil/Gas by Unit

Description	Ratio of Fischer-Tropsch to Oil/Gas	
	Fixed Capital Investment	FCI/BOE
Mining & Coal Preparation	0.82	1.05
Conversion	0.81	1.03
Process Gas Production	1.78	2.28
SNG Separation & Treatment	1.24	1.58
Product Finishing	0.90	1.15
Utilities	1.69	2.16
Environmental & General Facilities	1.04	1.33
Total	1.19	1.53

Table IV. Comparison of Projected Total Capital Requirements for Fischer-Tropsch and Oil/Gas Complexes

Item	Fischer-Tropsch $ MM	Oil/Gas $ MM	Ratio F-T - O/G
Fixed Capital Investment	1550	1300	1.19
Initial Catalyst & Chemicals	11	9	1.22
Start-Up Costs	110	86	1.28
Construction Financing[a]	212	188	1.13
Working Capital	113	107	1.06
Land, Rights of Way	1	1	1.00
TOTAL	1997	1691	1.18
Say	2000	1700	

a) Example: For 65/35 debt/equity, 9% interest, 0.75% commitment fee case.

Table V. Comparison of Projected Annual Operating Costs for Fischer-Tropsch and Oil/Gas Complexes

Cost Center	Annual Operating Costs - $MM	
	Fischer-Tropsch	Oil/Gas
Coal Mine	84.5	104.2
Coal Prepartion	2.3	3.2
Process Plant	101.5	84.4
Power Plant	7.7	--
Offsites	7.8	14.4
TOTAL	203.8	206.2
Say	205	205

Table VI. Comparison of Projected Average Required Product Selling Price at 12% DCF

Project Financial Structure	Required Average Product Selling Price in Dollars per Million BTU		
	Fischer-Tropsch	Oil/Gas	Ratio F-T - O/G
100% Equity	3.30	2.50	1.32
65/35 Debt/Equity	2.55	1.95	1.28
Breakeven	1.50	1.20	1.20

Table VII. Sensitivities of Average Required Product Selling Price to Key Economic Parameters

Economic Parameter	Sensitivity of Average RPSD, %			
	Fischer-Tropsch		Oil/Gas	
	100% Equity	65% Debt	100% Equity	65% Debt
Fixed Capital Investment	87	81	82	78
Operating Costs	15	19	21	27
Run of Mine Coal Costs	21	25	30	34

Table VIII. Possible Product Sales Values for Fischer-Tropsch Complex

Product	Daily Production	Possible Unit Sales Value in Dollars	Annual Gross Revenue in $ Million
SNG	260.0 MMscfd	4.25/Mcf	362.8
Liquids			
C_4s	3,535 BPD	12.00/bbl	14.0
Naphthas			
Light	10,620 BPD	15.50/bbl	54.3
Heavy	9,555 BPD	17.00/bbl	53.6
Alcohols	3,910 BPD	25.00/bbl	32.3
Diesel Fuel	16,960 BPD	14.50/bbl	79.9
Premium Fuel Oil	4,960 BPD	15.00/bbl	24.5
			241.6
Power	3,352 MW/hr	0.03/kW-hr	33.2
Total Energy			651.6
Sulfur	1,015 Ton	60/ton	20.1
Total 4th Qtr. 1975			671.7
Escalation (9% from 4th Qtr. 1975 to 4th Qtr. 1976)			60.5
Total 4th Qtr. 1976			732.2

Table IX. Possible Product Sales Values for Oil/Gas Complex

Product	Daily Production	Possible Unit Sales Value in Dollars	Annual Gross Revenue in $ Million
SNG	170 MMscfd	4.25/Mcf	238.425
Propane	6,030 BPD	11.00/bbl	21.890
Butane	4,100 BPD	12.00/bbl	16.235
Naphthas	9,400 BPD	15.50/bbl	48.080
Fuel Oil	56,400 BPD	9.75/bbl	181.470
Total Energy			506.100
Sulfur	118 LT/D	60/ton	2.335
Ammonia	90 ST/D	120/ton	5.565
Total 4th Qtr. 1975			514.000
Escalation (9% from 4th Qtr. 1975 to 4th Qtr. 1976)			46.000
Total 4th Qtr. 1976			560.000

Table X. DCF's for Possible Product Revenues

Project Financial Structure	DCF	
	Fischer-Tropsch	Oil/Gas
100% Equity	17	13
65/35 Debt/Equity	27	20

260

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The Analysis of Finned Catalytic Heat Exchangers

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The use of finned heat transfer surfaces has been suggested as a means of providing adequate catalytic surface and heat transfer surface in catalytic reactors. In these devices, temperature control is essential in order to maximize the yield and to insure long catalyst life. A thermal analysis of fins with a catalytic reaction occurring on their surface is made. It is shown that the existing solutions for conventional fins can be adapted to this case by defining an effective temperature for the reacting gas which includes the heat of reaction. This result is applied to the analysis of simple catalytic reactor/heat exchangers. Equations are developed which are useful in the design of this type of heat exchanger. The methods are illustrated by means of an example.

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The Analysis of Finned Catalytic Heat Exchangers

J. P. CALLINAN **D. L. BURFORD**

ABSTRACT

The use of finned heat transfer surfaces has been suggested as a means of providing adequate catalytic surface and heat transfer surface in catalytic reactors. In these devices, temperature control is essential in order to maximize the yield and to insure long catalyst life. A thermal analysis of fins with a catalytic reaction occurring on their surface is made. It is shown that the existing solutions for conventional fins can be adapted to this case by defining an effective temperature for the reacting gas which includes the heat of reaction. This result is applied to the analysis of simple catalytic reactor/heat exchangers. Equations are developed which are useful in the design of this type of heat exchanger. The methods are illustrated by means of an example.

NOMENCLATURE

A	fin cross sectional area
B	hL/k
C	$\dot{m}c_p$
c_p	specific heat capacity
f_1	A/L^2
f_2	P/L
h	Convective heat transfer coefficient
k	thermal conductivity
L	fin length
m	$(hP/kA)^{1/2}$
$\dot{m}$	mass flow rate
P	perimeter
$\dot{Q}$	heat of reaction per unit surface area
q	rate of heat transfer
r_f	fouling resistance
S	total surface area
s	surface area variable
T	temperature
T^*	$T + \dot{Q}/h_c$
t	thickness
x	fin length variable
α	defined by equation (35)
β	defined by equation (32)
γ	C_∞/C_i
δ	defined by equation (33)
ϵ	S_c/S_o
η_f	fin efficiency
θ	defined by equation (7) or equation (30)
ξ	defined by equation (6)
χ	defined by equation (31)

Subscripts

b	bare (unfinned) portion of tube
c	catalytic
cv	convection
f	fin
i	fluid flowing inside of tube (coolant)
k	conduction
L	at $x = L$
m	tube metal
o	outside
r	reaction
u	uncoated
∞	fluid flowing over finned tubes (reactant gas)
1	beginning of heat exchanger (where reactant gas enters)
2	end of heat exchanger (where reactant gas leaves)

INTRODUCTION

The current interest in the synthesis of clean fuels has resulted in the re-examination of previously developed chemical processes and in the development of new chemical processes for the production of synthetic fuels. Many of these processes employ exothermic shift reactions, such as the water gas shift or methanation, which are promoted by the use of catalysts [1, 2 and 3]*. The yield of these reactions is strongly dependent upon temperature [4]. Further, damage to the catalyst can occur if excessive temperatures are achieved [5].

Using conventional catalytic bed reactors for exothermic reactions necessitates the use of multistage reactors with intermediate heat exchangers in order to control the reaction temperature, thus maximizing the yield [5]. Providing a single surface which combines the functions of catalysis and heat exchange can result in improved temperature control. Wei and Chen [3] discuss the use of tubular heat exchangers in which one side of the tubes is coated with a catalyst. The reactant gas

*Numbers in brackets refer to references listed at the end of the paper.

flows on this side of the tubes while a coolant flows on the opposite side. The addition of fins to the catalyst side of the heat exchanger can decrease the volume and cost of such "catalytic heat exchangers" while still achieving adequate temperature control [7].

Typical finned tubes which could be employed in finned catalytic heat exchangers are illustrated in Figure 1. The catalyst would be deposited (e.g., flame sprayed [3]) on all or part of the finned side of the tube. Variations (e.g., the entire finned side of the tube coated with the catalyst, alternate fins coated or fins coated on one side only) would permit designing the heat exchanger to achieve optimal temperatures. The reactant gas flows over the finned surface while the coolant (possibly a phase change substance) would flow through the inside of the tubes.

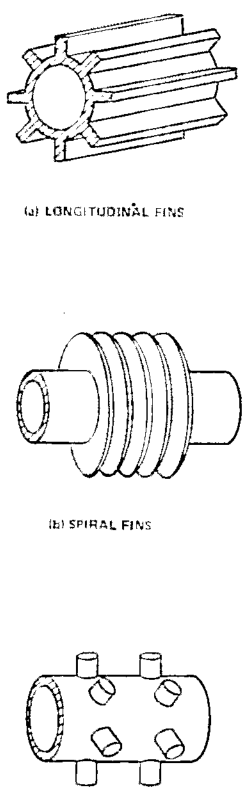

Fig. 1 Typical finned-tube geometries

The work which follows is a thermal analysis of fins on whose surface a chemical reaction is occurring and a thermal analysis of simple heat exchangers employing finned, catalytic surfaces.

ANALYSIS OF CATALYTIC FIN

A fin of arbitrary geometry is shown in Figure 2. In analyzing this fin, the usual fin assumptions [6, pp. 85 and 86] are made. In addition, it is assumed that an exothermic chemical reaction is occurring on the surface of the fin, releasing energy at the surface [3]. A heat balance on the surface area element $P(x)dx$ yields: the heat convected from the surface plus the heat conducted from the surface equals the heat released at the surface due to the chemical reaction. The heat convected from the surface is:

$$dq_{cv} = h_c(T - T_\infty)\; P(x)dx \qquad (1)$$

where the convective heat transfer coefficient, h_c, is assumed constant and includes the effect of mass transfer on heat transfer. Thermal radiation is neglected in this analysis although its effect could be incorporated in h_c.

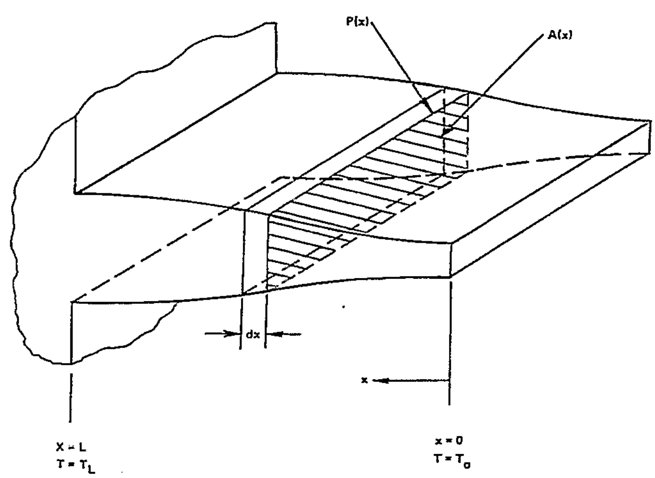

Fig. 2 Fin of arbitrary geometry

The heat conducted from the surface equals the net heat conducted out of the volume element $A(x)dx$,

$$dq_k = -k \frac{d}{dx}\left[A(x)\frac{dT}{dx}\right] dx \qquad (2)$$

The energy released by the chemical reaction occurring at the surface is

$$dq_r = \dot{Q}\; P(x)dx \qquad (3)$$

Combining equations (1), (2) and (3) to form a heat balance at the surface yields

$$h_c(T - T_\infty)P(x)dx - k\frac{d}{dx}\left[A(x)\frac{dt}{dx}\right] dx = \dot{Q}P(x)dx \qquad (4)$$

This can be rearranged as follows

$$A(x)\frac{d^2T}{dx^2}+\frac{dT}{dx}\frac{dA(x)}{dx}-\frac{h_cP(x)}{k}\left[T-\left(T_\infty+\frac{\dot{Q}}{h_c}\right)\right]=0 \tag{5}$$

Defining the following dimensionless variables,

$$\xi = x/L \tag{6}$$

$$\theta = \frac{[T-(T_\infty+\dot{Q}/h_c)]}{[T_L-(T_\infty+\dot{Q}/h_c)]} \tag{7}$$

$$f_1(\xi) = A(L\xi)/L^2 \tag{8}$$

$$f_2(\xi) = P(L\xi)/L \tag{9}$$

$$B = h_cL/K, \tag{10}$$

equation (5) becomes

$$f_1(\xi)\frac{d^2\theta}{d\xi^2}+\frac{d\theta}{d\xi}\frac{df_1(\xi)}{d\xi} - f_2(\xi)B = 0 \tag{11}$$

Typical boundary conditions are:

$$\text{at } \xi = 0,\ (a)\ \frac{d\theta}{d\xi} = 0 \quad \text{(adiabatic tip)} \tag{12-a}$$

$$\text{or (b) } \frac{d\theta}{d\xi} = B\theta \quad \text{(convective tip)} \tag{12-b}$$

$$\text{and at } \xi = 1,\ \theta = 1 \quad \text{(base temperature is known)} \tag{13}$$

Equation (11) with its boundary conditions, equations (12) and (13), is identical to the heat conduction equation for a conventional fin of arbitrary geometry (Ref. 6, Chapter 2). It follows that published solutions to this equation for conventional fins of various specific geometries are also applicable to the catalytic fins of identical geometry discussed here. In applying these existing solutions to catalytic fins, one simply replaces the fluid temperature, T_∞, with an effective fluid temperature, $T_\infty^\star$, defined as

$$T_\infty^\star = T_\infty + \dot{Q}/h_c \tag{14}$$

where $T_\infty^\star$ is assumed to be constant.

In general, the heat transfer through the base of the fin is given by

$$q_f = -hS[T_L - T_\infty^\star]\eta_f \tag{15}$$

and the heat transfer to the fluid is given by

$$q_\infty = \dot{Q}S - q_f \tag{16}$$

Where S is the surface area of the fin.

Fin efficiencies, η_f, for specific fin geometries appear in the heat transfer literature (e.g., Ref. 6, pp. 97, 110, 123 and 135 through 162).

As an example, consider the longitudinal fin of constant profile (Figure 1-a) with an adiabatic tip. From the wellknown solution for the conventional fin of this geometry, the fin temperature distribution is

$$\theta = \frac{\cosh[mL(1-\xi)]}{\cosh mL} \tag{17}$$

$$\text{where } m^2 = h_cP/kA \tag{18}$$

P = perimeter (constant)

A = cross sectional area (constant)

And, the rate of heat transfer through the base of the fin is given by equation (15) where

$$\eta_f = \frac{\tanh(mL)}{mL} \tag{19}$$

ANALYSIS OF CATALYTIC HEAT EXCHANGER

A section of a generalized heat exchanger tube is shown in Figure 3. In this heat exchanger some of the finned area, S_{fc}, has a catalytic coating and the remainder, S_{fu}, doesn't. Also, a portion of the outside (unfinned) surface area, S_{bc}, of the tube is coated and the remainder, S_{bu}, is not. The heat flow to the coolant in the tube can be divided into two parallel paths, that coming from the catalytic portion of the tube outer surface, q_{ic}, and that coming from the reactant gas through the uncoated surface, q_{iu}.

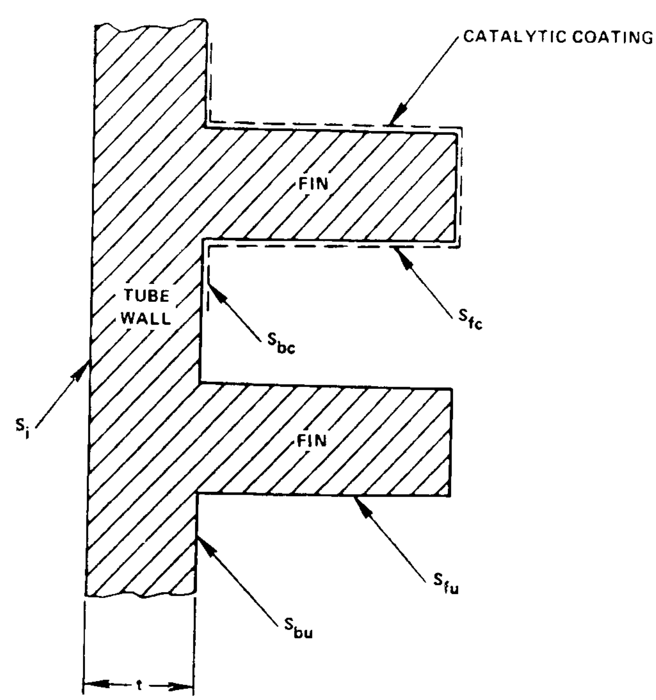

Fig. 3 Section of generalized catalytic heat exchanger tube

By making a heat balance on a differential element of the surface where the catalytic reaction is occurring (similar to what was done in the previous section) it can be shown that

$$dq_{ic} = U_c\left(T_\infty + \frac{\dot{Q}}{h_c} - T_i\right)ds_o \tag{20}$$

where

$$\frac{1}{U_c} = \frac{S_o}{h_i S_{ic}} + r_{fi}\frac{S_o}{S_{ic}} + \frac{tS_o}{kS_{mc}}$$

$$+ \frac{S_o}{h_c(S_{bc} + \eta_f S_{fc})} \tag{21}$$

For the uncoated portion of the tube,

$$dq_{iu} = U_u(T_\infty - T_i)\ ds_o \tag{22}$$

where

$$\frac{1}{U_u} = \frac{S_o}{h_i S_{iu}} + r\ \frac{S_o}{S_{iu}} + \frac{tS_o}{kS_{mu}}$$

$$+ \frac{S_o}{h_u(S_{bu} + \eta_f S_{fu})} \tag{23}$$

The total heat transfer to the coolant is, from equations (20) and (22),

$$dq_i = dq_{ic} + dq_{iu}$$

$$= U(T_\infty - T_i)ds_o + \frac{U_c}{h_c}\dot{Q}ds_o \tag{24}$$

where $U = U_u + U_c$

Since the only heat source is the catalytic reaction, the heat transfer to the reactant gas is

$$dq_\infty = \dot{Q}\frac{S_c}{S_o}ds_o - dq_i$$

$$= \dot{Q}\left(\frac{S_c}{S_o} - \frac{U_c}{h_c}\right)ds_o$$

$$- U\ (T_\infty - T_i)\ ds_o \tag{25}$$

For this differential element of the heat exchanger the coolant temperature, T_i, increases by an amount*

$$dT_i = dq_i/C_i \tag{26}$$

and the temperature of the reactant gas, T_∞, increases by an amount*

$$dT_\infty = dq_\infty/C_\infty \tag{27}$$

Combining equations (26) and (27),

$$dT_\infty - dT_i = dq_\infty/C_\infty - dq_i/C_i. \tag{28}$$

Eliminating dq_∞ and dq_i with equations (24) and (25),

$$dT_\infty - dT_i = \dot{Q}\left[\frac{S_c}{S_o}\frac{1}{C_\infty} - \left(\frac{1}{C_\infty} + \frac{1}{C_i}\right)\frac{U_c}{h_c}\right]ds_o$$

$$-\left(\frac{1}{C_\infty} + \frac{1}{C_i}\right)U\ (T\ \ - T_i)\ ds_o \tag{29}$$

Defining the following dimensionless variables

$$\theta = \frac{T_\infty - T_i}{T_{\infty 1} - T_{i1}} \tag{30}$$

Where the subscript 1 refers to the point where $s_o = 0$ (inlet conditions for a parallel flow heat exchanger),

$$\chi = s_o/S_o \tag{31}$$

$$\beta = \frac{\dot{Q}S_o}{C_\infty\ (T_{\infty 1} - T_{i1})} \tag{32}$$

$$\delta = \left[\varepsilon - (1 + \gamma)\ \frac{U_c}{h_c}\right] \tag{33}$$

$$\gamma = C_\infty/C_i \tag{34}$$

$$\varepsilon = S_c/S_o$$

and $$\alpha = (1 + \gamma)\ \frac{US_o}{C_\infty} \tag{35}$$

equation (31) becomes

$$d\theta = \beta\delta d\chi - \alpha\theta d\chi \tag{36}$$

Rearranging equation (29)

$$\frac{d\theta}{d\chi} + \alpha\theta = \beta\delta \tag{37}$$

where the boundary condition is at $\chi = 0$, $\theta = 1$. Solving equation (37) and eliminating the constant of integration with the boundary condition, the relationship for the temperature difference in the heat exchanger is

$$\theta = \left(1 - \frac{\beta\delta}{\alpha}\right)e^{-\alpha\chi} + \frac{\beta\delta}{\alpha} \tag{38}$$

For the entire heat exchanger (i.e., at $\chi = 1$),

$$\theta_2 = \left(1 - \frac{\beta\delta}{\alpha}\right)e^{-\alpha} + \frac{\beta\delta}{\alpha} \tag{39}$$

Defining

$$\theta_i = \frac{T_i\ \ - T_{i1}}{T_{\infty 1} - T_{i1}} \tag{40}$$

and $$\theta_\infty = \frac{T_\infty\ \ - T_{\infty 1}}{T_{\infty 1} - T_{i1}} \tag{41}$$

---

*Here we are assuming a parallel flow heat exchanger in which the temperature of both fluids are assumed to increase as s_o increases

We note that

$$\theta_\infty - \theta_i = \theta - 1 \tag{42}$$

and further, that

$$\dot{Q}S_c\chi = C_\infty(T_\infty - T_{\infty 1}) + C_i(T_i - T_{i1})$$

or,

$$\epsilon\beta\chi = \theta_\infty + \theta_i/\gamma \tag{43}$$

Combining equations (38), (42) and (43) we obtain expressions for the dimensionless temperatures of each fluid. For the reactant gas

$$\theta_\infty = \left(\frac{1}{1+\gamma}\right)\left[\gamma\epsilon\beta\chi + \left(\frac{\beta\delta}{\alpha} - 1\right)\left(1 - e^{-\alpha\chi}\right)\right] \tag{44}$$

and for the coolant

$$\theta_i = \left(\frac{\gamma}{\gamma+1}\right)\left[\epsilon\beta\chi - \left(\frac{\beta\delta}{\alpha} - 1\right)\left(1 - e^{-\alpha\chi}\right)\right] \tag{45}$$

In the design of a catalytic heat exchanger the objective is to produce a specified yield of a particular product of the reaction. The chemical kinetics would specify the reactant gas flow rate, the total surface area of catalyst required and the allowable range of reactant gas temperature, T_∞. Equations (39), (44) and (45) can then be used to size a heat exchanger which will accomplish these objectives. These same equations apply equally well to the counterflow case and to the case when the coolant undergoes a phase change. For the counterflow case C_i must be defined as follows,

$$C_i = -\dot{m}_i \, c_{pi} \tag{46}$$

and the coolant enters the heat exchanger at T_{i2}. For the case in which the coolant undergoes a phase change (i.e., T_i is constant),

$$\gamma = 0 \tag{47}$$

and equation (45) has no meaning.

The expression for the temperature difference between the reactant gas and the coolant, equation (38), consists of two terms, an exponential decay term and a constant. As the reactant gas flows through the heat exchanger its temperature changes rapidly at first. As $\alpha\chi$ gets very large compared to one, T_∞ remains at a fixed increment greater than the coolant temperature. This can be seen by evaluating equation (38) for $\alpha\chi >> 1$, where θ takes on its asymptotic value,

$$\theta = \frac{\beta\delta}{\alpha} \tag{48}$$

Using equations (30), (32), (33) and (34) the asymptotic value can be expressed in a dimensional form.

$$T_\infty - T_i = \frac{\dot{Q}}{U}\left[\frac{S_c}{S_o}\left(\frac{1}{1+\gamma}\right) - \frac{U_c}{h_c}\right] \tag{49}$$

This is illustrated in Figures 4 and 5. Figure 4 shows the effect of $\beta\delta/\alpha$ on the temperature difference for $\alpha = 20$. The three cases shown represent the initial reactant gas temperature difference being lower than, equal to and greater than $\beta\delta/\alpha$, the asymptotic value. In Figure 5 it is seen that the temperature difference achieves the asymptotic value within the first 5% of the length for $\alpha > 100$.

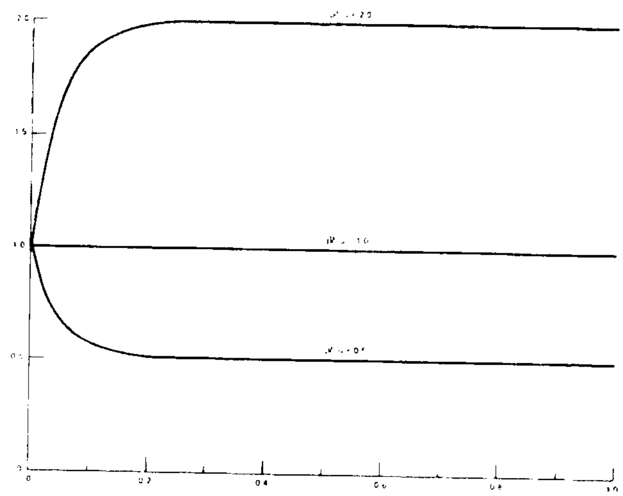

Fig. 4 Temperature difference versus length for $\alpha = 20$

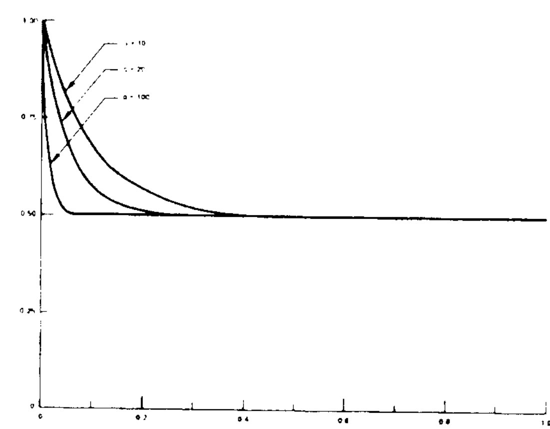

Fig. 5 Temperature difference versus length for $\beta\delta/\alpha = 0.5$

In cases involving highly exothermic reactions, $\alpha >> 10$ and, therefore, $\alpha\chi >> 1$ occurs for small values of χ. Equation (49) states that, under this condition, the reactant gas temperature will remain a constant increment above the coolant temperature. If the coolant temperature remains constant (phase change case) the reactant gas temperature will also remain constant. Equation (49) shows quantitatively the effect of design parameters on T_∞. For specified values of $\dot{Q}$ and S_c, the value of T_∞ can be reduced by increasing U, by increasing S_o, by increasing U_c or by decreasing h_c. It should be noted that these parameters are interrelated.

The effect of γ, the ratio heat capacity rate for the reactant gas to that of the coolant, is shown in Figure 6 for $\beta\delta/\alpha = 0.5$ (a parallel flow configuration). As one would expect, the coolant temperature increase more for the larger value of γ.

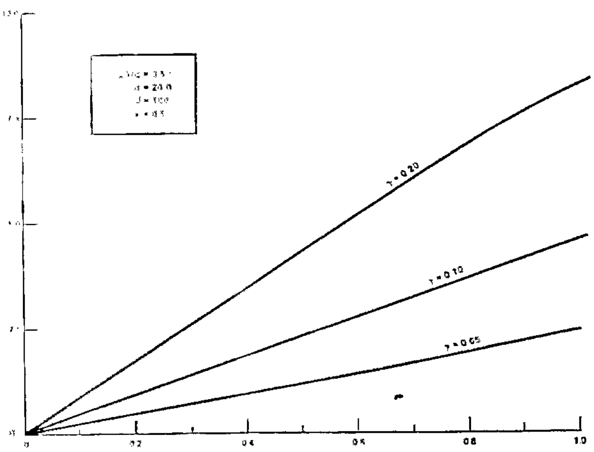

Fig. 6 Coolant temperature versus length

EXAMPLE

The following example illustrates the application of the analytical techniques developed above and also illustrates quantitatively the nature of the temperature variations in this type of heat exchanger. The data are representative of processes which occur in coal gasification technology.

It is desired to cause 6 kg/s of a reactant gas to undergo a particular shift reaction. 4500 m^2 of catalyst surface are required. The reactant gas enters the heat exchanger at 295°C. It is desired to maintain the gas temperature in a range between 310 and 330°C in order to maximize the yield of the reaction. A heat exchanger utilizing spiral fins (Figure 1-b) is selected. After some preliminary study a heat exchanger design is selected. The appropriate data for the design selected is summarized in Table 1. Three coolant schemes will be examined, a phase change coolant, a sensible heat coolant in a parallelflow configuration and a sensible heat coolant in a counterflow configuration. The appropriate data and calculated parameters for each coolant scheme is summarized in Table 2.

The temperature distributions for this heat exchanger with the three coolant schemes are shown in Figures 7, 8 and 9. Only the phase change coolant scheme maintains the temperature of the reactant gas in the specified range. It should be noted that in the counterflow case, the coolant inlet temperature is T_{i2} and T_{i1} can be calculated from equation (49). β can then be calculated using equations (32), (41) and (44).

TABLE 1. Data for Catalytic Heat Exchanger Design

S_o (m^2)	9000
S_c (m^2)	4500
S_o/S_i	10.2
$\dot{m}_\infty$ (kg/s)	6
$T_{\infty 1}$ (°C)	295
$c_{p\infty}$ (J/kg°C)	2500
C_∞ (W/°C)	15,000
h_c (W/m^2°C)	700
h_i (W/m^2°C)	8000
r_{fi} (m^2°C/W)	0.00009
U_c (W/m^2°C)	100
U_i (W/m^2°C)	100
U (W/m^2°C)	200
$\dot{Q}$ (W/m^2)	10,000

TABLE 2. Data for Specific Coolant Schemes

	Phase Change	Parallel-Flow	Counter-Flow
Coolant Inlet Temp. (°C)	305	235	235
$\dot{m}_i$ (kg/sec)	-	120	120
c_{pi} (J/kg°C)	-	4167	4167
C_i (W/°C)	-	500,000	-500,000
γ	0	0.03	-0.03
ε	0.5	0.5	0.5
δ	0.357	0.353	0.361
α	120	124	116
β	-600	100	-192
$\beta\delta/\alpha$	-1.785	.285	-.598

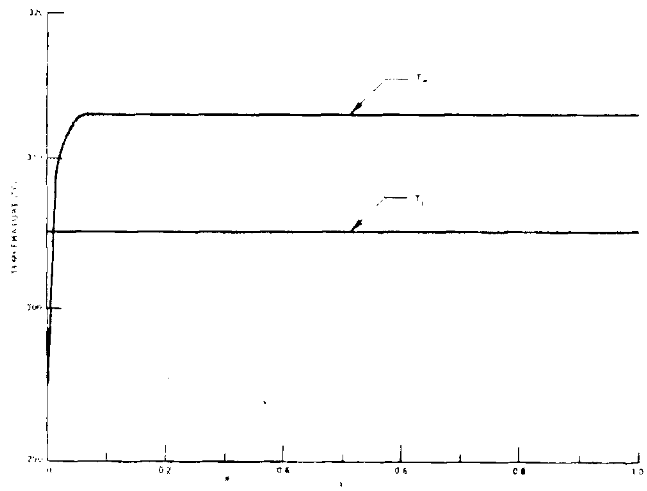

Fig. 7 Heat exchanger temperature distribution. phase-change coolant scheme ($\gamma = 0$)

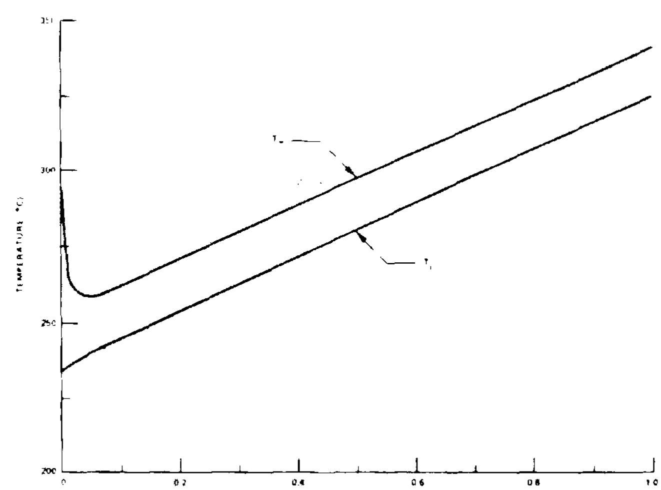

Fig. 8 Heat exchanger temperature distribution, parallel-flow coolant scheme

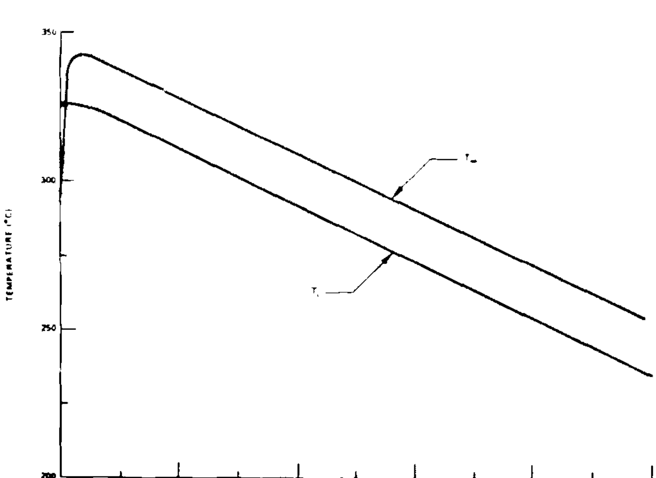

Fig. 9 Heat exchanger temperature distribution counterflow coolant scheme

DISCUSSION

An examination of the results of the above example reveal some of the unique features of this type of heat exchanger. Temperature crossover is possible. The temperature of the hot fluid can be reduced by increasing the thermal resistance of the hot fluid side, $1/h_c$, in equation (49).

The equations developed in this paper can be used for the thermal design of a component which performs the dual functions of a catalytic chemical reactor and a heat exchanger. The type of catalytic reactor-heat exchanger analyzed has the potential of providing excellent control of the temperature of the reacting gas. The ideal design would employ a phase change coolant having $\beta\delta/\alpha = 1.00$. Under these conditions, according to this simplified theory, the temperature of the reactant gas would remain constant throughout the heat exchanger. If it were not possible to design the exchanger such that $\beta\delta/\alpha = 1$ then a large value of α is desirable.

Degradation of the catalyst during the operating life of the heat exchanger would result in the value of β decreasing. This would result in a reduction in the asymptotic temperature below the design value. This could be compensated for by decreasing the flow of reactant gas as can be seen in equation (32).

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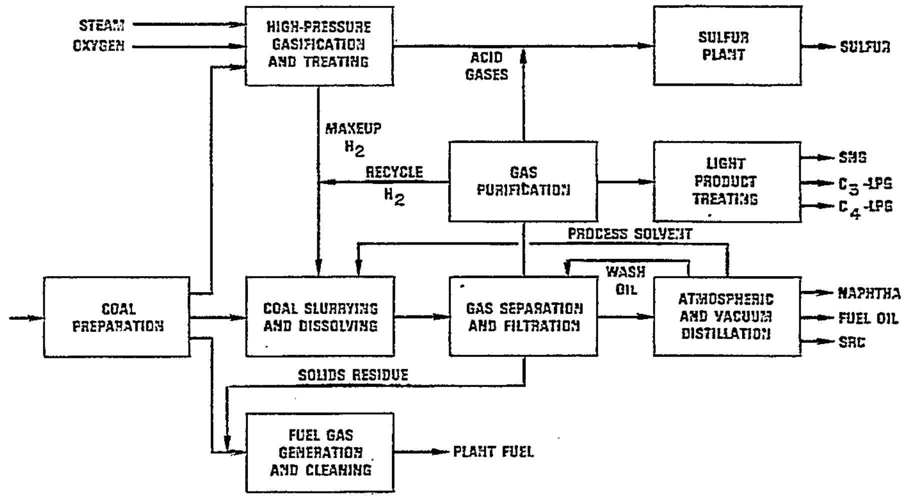

Figure 1. Simplified block flow diagram of the oil/gas plant.

COAL PROCESSING

Oil/Gas Plant Design Criteria

This coal conversion facility will feed 50,000 ton/day to an SRC based hydroliquefaction unit and a pressure entrained gasifier.

J. B. O'Hara, G. H. Hervey, S. M. Fass, and E. A. Mills
The Ralph M. Parsons Co., Pasadena, Calif.

A conceptual design and economic evaluation for a commercial scale oil/gas coal conversion complex, being developed by Ralph M. Parsons Co., is one of several designs to be completed under contract to the Energy Research and Development Administration (ERDA).

The oil/gas concept recognizes that methane and other light hydrocarbons are produced both in coal liquefaction and gasification. It uses appropriate gasification and liquefaction products to produce significant amounts of substitute natural gas (SNG).

Earlier preliminary conceptual design/economic assessment work performed for Project Independence envisioned a complex to produce 100,000 bbl./day of liquid fuel plus significant SNG. Results were summarized in a presentation to the Project Independence Blueprint hearings. (*1*) The effort reported here is an extension of that work.

Parsons is actively assisting ERDA to develop viable commercial plants for the conversion of coal to clean fuels. There are two distinct elements involved in this role:

1. Preliminary design services in which Parsons develops preliminary conceptual designs and economic evaluations for commercial coal conversion plants.

2. Parsons also supplies technical evaluation contractor services to assist ERDA in monitoring certain of the liquefaction development programs.

A simplified block flow diagram depicting the process concept is shown in Figure 1. The concept is based on:

1. Production of liquid and/or de-ashed, low sulfur solid fuels by reaction of hydrogen with coal.

2. Production of SNG from the following *potential* sources: 1) methane produced in the hydroliquefaction step; 2) methane produced in the gasification step; 3) steam reforming of LPG followed by processing to produce a methane-rich stream.

Predesign studies

A program was completed to provide a quantitative economic basis for selection of the preferred process configuration. Procedures used began with process flow diagrams with heat and material balances plus equipment requirements adequate to estimate fixed capital investments. These were developed for the separate alternatives.

Operating costs were estimated and an economic comparison of the alternatives using discounted cash flow rate of return procedures was then made. The results of the comparison were reviewed, and the preferred option selected.

Summaries of the results of process preference studies are shown in Table 1. This data has been incorporated into the design criteria.

The design criteria are intended to 1) describe key elements of the design that will be created to permit users to anticipate size, product slate, and general characteristics of the resulting facility, and 2) permit designers to proceed with their objectives and work.

The plant would be located in the Eastern Interior (coal) Region of the U.S. *Coal feed* would be Illinois No. 6 seam coal, captive mine; approximately 40,000 tons/day of cleaned, sized coal. *Plant design* incorporates coal gasification in an entrained gasifier. Coal dissolving will be based on the SRC process. The design will include preferred options for this plant as determined from the results of process preference studies. An oil/gas B.t.u. value ratio of approximately 2 is expected. The technology offers flexibility; ratios as high as 6 have been studied.

Products include 170 million std. cu. ft./day of SNG plus LPG, 11,000 bbl./day of LPG; naphtha, 9,000 bbl./day; fuel oil, 60,000 bbl./day; sulfur, 1,400 ton/day; and ammonia, 50 ton/day.

Probable configuration

The probable configuration of the complex is outlined below.

Coal mine: approximately 50,000 ton/day of run-of-mine (ROM) coal will be produced. Five separate mining units are planned.

Coal preparation: this plant is designed to receive 50,000 ton/day of ROM coal and produce 40,000 ton/day of feed for plant use. This coal is dried and reduced to the appropriate size for use in the gasifiers and the dissolver.

Dissolving and filtration: coal feed to the dissolving unit is slurried with recycle solvent, hydrogen-containing gas is added, and the mixture fed to the preheater and dissolver at elevated temperature and pressure. Product slurry will be separated from gases, and then flashed in various stages to remove light constituents from the filter feed slurry. Filters will separate solids from the remaining liquid. The filter cake will be washed with a light solvent to remove most of the adhering liquids, and then discharged to a dryer to recover wash solvent. Filtrate will proceed to distillation and the dry filter cake to fuel gas production.

High pressure gasification: fresh coal will be fed to a pressure entrainment-type gasifier; the product gases are fed to a sour shift unit to convert most of the carbon monoxide to hydrogen. A selective acid gas removal unit will be used to separate carbon dioxide and hydrogen sulfide into two streams; the first for disposal and the other for sulfur plant feed.

Distillation: liquids produced in the dissolving section plus plant condensate streams are separated by distillation into the required recycle and product streams.

Gas treating: off-gas from the dissolver will be treated to remove carbon dioxide and hydrogen sulfide, and the sweet gas will be sent to a cryogenic separation unit to remove methane and heavier hydrocarbons. The hydrogen and carbon monoxide will be recycled to the dissolver preheater. The methane will be separated from the LPGs, and purification units will produce the specification SNG and LPG.

Fuel gas production: dried filter cake and additional fresh coal will be fed to a low pressure, air-fed, entrained gasifier to produce low B.t.u. gas for plant fuel needs.

Water and effluent gas treating: all plant water streams will be treated. The gases, consisting mainly of ammonia and hydrogen sulfide, will be separated to produce anhydrous ammonia, a saleable product, and hydrogen sulfide for feed to the sulfur plant. The sulfur plant will in turn convert the hydrogen sulfide to elemental sulfur and a clean stack gas.

Acknowledgments

Preparation of a design and economic evaluation of this nature requires the contribution, support, and guidance of a number of organizations and people. The contributions of the Energy Research and Development Administration-Fossil Energy, and the contributions of the many people in Parsons are gratefully acknowledged. #

Table 1. Oil/gas plant process preference study results (study description in italics).

Hydrogen vis-a-vis *Syngas as hydroliquefaction agent:*
Syngas shows slight reduction (i.e., ≈ 3%) in required product selling price.

Procedure for separation of NH_3 from NH_3 - H_2S mixtures:
Separation by fractionation shows lower fixed capital investment and lower operating costs. Annual estimated savings are about $400,000/yr., equivalent to about 0.25% reduction in total base annual revenue requirement.

Dissolver residence time:
Reduction of nominal liquid space time from 60 to 30 min. is estimated to reduce the required annual revenue by $2.8 million/yr., equivalent to about 1.7% reduction in total base annual revenue requirement.

Production of incremental SNG by reforming LPG:
Reforming LPG to produce SNG increases required annual revenue by about $1 million/yr., equivalent to about 0.6% of total base annual revenue requirement.

Acid gas removal procedure:
Physical solvent showed required annual revenue reduction of about $5 million *vis-a-vis* chemical absorbent; or approximately 3% of total base annual revenue requirement.

Sour vis-a-vis *sweet shift:*
Sour shift shows about $6.2 million reduction in required annual revenue; or approximately 3.8% of total base annual revenue requirement.

Use of power recovery turbines:
Use of turbines will reduce required annual revenues by $400,000 at 30 mills/kW-hr. rate; this is about 0.3% of total base annual revenue requirement.

Recovery of liquids from filter cake:
Addition of this operation is estimated to reduce the required product selling price, in $/million B.t.u., by about 14%.

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OIL/GAS COMPLEX
CONCEPTUAL DESIGN/ECONOMIC ANALYSIS

OIL AND SNG PRODUCTION

R & D REPORT NO. 114 - INTERIM REPORT NO. 4

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SECTION 1

INTRODUCTION

This report presents the results of a conceptual design and economic evaluation for a commercial complex to mine high-sulfur coal and produce substitute natural gas (SNG), fuel oil, naphtha, and liquefied petroleum gases (LPG) using hydroliquefaction technology for the coal conversion portion of the complex.

This work was performed for the Energy Research and Development Administration (ERDA) - Fossil Energy, Demonstration Plants Division, whose guidance and support in these activities are gratefully acknowledged. The design uses the teachings of the ERDA-sponsored solvent refined coal (SRC) hydroliquefaction and entrained slagging gasification programs, with adaptation to the specific Oil/Gas objectives. Pseudo catalytic SRC II hydroliquefaction techniques are used in which a portion of the hydroliquefier effluent is recycled to the hydroliquefier reactor to provide a higher content of ash constituents, longer reaction time, and greater hydrogen consumption to produce products that are primarily gases and liquids at ambient conditions.

The design basis was developed in cooperation with ERDA.

1.1 OBJECTIVES

The objectives of the work described in this report are to:

- Develop a conceptual design for a commercial grass roots hydroliquefaction-based industrial complex including all operations required to mine coal, prepare it by cleaning and washing it, and convert it to ecologically clean liquid and gaseous fuel products. The design should be capable of producing fuels at a price competitive with alternative sources.

- Define the projected product characteristics and marketability.

- Define probable project and financial parameters for design, engineering, procurement, construction, and startup of the complex.

- Estimate the economics for the facility to serve as a guide in making decisions regarding future commercial applications of this technology.

- Provide recommendations regarding additional development effort to foster commercial exploitation of the technology.

1.2 REPORT ORGANIZATION

A summary of the various parts of this report is presented in Section 2 to aid in rapid assimilation of its contents.

Sections 3 and 4 provide an introduction and orientation for the detailed design information presented in later sections. Design parameters and design bases used are summarized in Section 3. Section 4 describes the project scope and major units included in the complex. The relationship of the major operational steps and material flows is presented in the form of a block flow diagram; a plot plan, an artist's rendition of the complex, and a photograph of a model of the complex are also included. Section 5 contains detailed descriptions of the separate units that make up the complex. The process flow diagrams are presented in Section 6.

Sections 7 through 10 present key process efficiency factors and product characteristics/marketability projections. Section 7 summarizes the material balance, and Section 8 presents the projected product characteristics and marketability. The energy balance is given in Section 9, and Section 10 is a detailed utility summary.

A detailed analysis of environmental factors is presented in Section 11. Flow diagrams showing the quantities and compositions of contaminant containing streams plus the facilities and treatments used to remove the contaminants are described in Section 11. Section 12 summarizes plant startup procedures. Section 13 summarizes the major equipment items required, and with the design information previously summarized in the report, provides the basis for the fixed capital investment and operating cost estimates that follow in Section 14. The estimated economics for the complex are developed in Section 14 where fixed capital investment, other capital requirements, operating requirements and operating costs, and projected profitability are presented, accompanied by pertinent sensitivity factors.

The experimental data used as a basis to design the key coal conversion steps are presented in Section 15. Process considerations and predesign studies completed to define the preferred process configuration are given in Section 16.

Finally, Sections 17 and 18 provide a retrospective review of the design. Section 17 presents judgments regarding the expected performance of the plant, and Section 18 points out further potential improvements.

SECTION 2

SUMMARY

A conceptual design and economic evaluation has been completed for a project to design, engineer, construct, start up, and operate an industrial complex to mine high-sulfur coal and convert it to SNG, LPGs, naphtha, and heavy fuel oil. The results are summarized in this report.

This work was done with the support and guidance of the ERDA - Fossil Energy, Demonstration Plants Division. The design basis, utilizing teachings from the SRC process development pfogram, was developed in cooperation with ERDA.

The scope of the industrial complex is a grassroots facility consisting of a large captive coal mine that produces approximately 47,000 tons per day (TPD) of run-of-mine (ROM) coal supplying the feed material to a coal preparation plant, which in turn supplies approximately 36,000 TPD of clean, washed coal to a hydroliquefaction-based coal conversion plant. In the facility, the feed coal is converted to the above-mentioned product slate; byproduct ammonia and sulfur are also produced. Low-Btu, low-sulfur fuel gas is produced as fuel for process furnaces and for a close-coupled steam and power generation plant that produces all utilities required for the captive use in the complex.

The complex is conceived to be located in the eastern region of the Interior Coal Province. The facility meets desired location criteria consisting of significant resource of high-sulfur coal with a large utility/industrial market nearby and with ecological restrictions on direct consumption of the indigenous high-sulfur coal.

Process flowsheets and accompanying heat and material balances are presented, based on a typical coal analysis that is intermediate between the extreme analyses that might be encountered during a 20-yr project life. The equipment was sized to handle this typical coal. The design provides for the simultaneous mining of five mine faces and the mixing of the resultant coals in a storage pile to produce a relatively uniform feed coal to the process plant.

Products from the process plant include the following approximate quantities:

- 56,000 bbl/day of fuel oil; characteristics are projected to be roughly equivalent to low-sulfur bunker C.
- 10,000 bbl/day of naphtha.
- 10,000 bbl/day of LPG (C_3 and C_4).

- 165 MM SCFD of SNG.
- 1,300 ST/D of sulfur.
- 90 ST/D of anhydrous ammonia.

The estimated fixed capital investment for the complex is $1.25 billion; all estimates are in fourth-quarter 1975 dollars. The total capital investment is estimated to be $1.4 billion, which includes the cost of initial raw materials, catalysts and chemicals, allowance for startup and land acquisition, and initial working capital.

The fixed capital investment estimate was independently evaluated by the U.S. Army Engineer Division (USAEDH), Huntsville, Alabama. This work was done under contract to ERDA, Contract No. EX-76-C-01-1759. The USAEDH estimate was approximately 4% lower than Parsons, and they report an indicated overall estimate confidence factor of ±4%.

A representative project schedule for design, engineering, construction, and startup is given; a 56-month schedule to mechanical completion is projected, and a probable fund drawdown schedule is presented.

ECONOMIC PROJECTIONS

The population of the complex is estimated to be about 2,350. Operating costs are projected to be about $195 million per year. The required plant revenue for a 12% discounted cash flow (DCF) rate of return with 65% debt at 9% interest is $395 million per year. The predicted required product selling price for these financial parameters is $1.80/MM Btu.

The design is considered to be workable with the understanding that the estimated costs that are reported here have the probability of being greater than if additional information were available, which is often the case for first-generation plants.

Predicted required product selling prices, expressed in dollars per million Btu, for 100% equity financing and a nonprofit (0% DCF) or breakeven boundary case in addition to the 65/35 debt equity case described previously are:

Financing Method	Selling Price, $
100% equity	2.35
Debt/equity ratio = 65/35	1.80
Break-even	1.15

These values correspond to approximately $10.80/bbl and $14.10/bbl of oil equivalent for the 65/35 debt/equity ratio and 100% equity cases, respectively; values are based on a heating value of 6 million Btu/bbl.

Using an arbitrary SNG selling price of $2.50 per M SCF would generate the required $395 million revenue with a liquid products sales value of $10.00/bbl of the 65/35 debt/equity case. Another alternative, using a recently published allowable price for coal-derived SNG of $4.20, will generate the required revenue with an average liquid products selling price of $6.50.

The sensitivities of required selling prices and profitability to key economic parameters are presented. The selling prices are highly capital sensitive.

The 3:1 solvent-to-coal ratio of feed to the coal dissolvers in Unit 12 in this design may be conservative. Recent pilot plant data indicates that a ratio as low as 1.5:1 could be used. This lower rate is a potential improvement and could reduce the fixed capital investment and required product selling prices by approximately 6% and 5%, respectively.

Methods of scale-up were carefully considered. The scale-up factor from the SRC pilot plant to this conceptual design was of the order of 400. However, the scale-up factor for the critical dissolver, which liquefies the coal by reaction with hydrogen, is approximately 135. The dissolver vessels specified are the largest that can be fabricated with existing materials, fabrication, and coding practices. Methods of scale-up were selected to provide efficiency, operability, and process control.

The design represents an assessment of a proposed configuration and potential economics for this type of technology. It projects a total thermal efficiency of approximately 77%, which means that more than three-quarters of the energy (Btu) contained in the feed coal is converted to low-sulfur fuel products. This efficiency is higher than predicted by earlier designs and is the result of detailed analysis of the efficiencies in all major plant units.

The design conceives operation in the mid-1980s and therefore proposes use of certain equipment and techniques that require further development prior to commercial operation. To accomplish this objective, the use of engineering judgment for the scale-up and the selection of the equipment was required. The design also represents an exposition of factors required to integrate the coal conversion plant with a large coal mine and a closely coupled electrical power generation facility for internal power requirements. The development of the interfaces between the coal mine, process plant, and power plant has defined a number of the design and operational options that exist to maximize efficiencies and profitabilities.

The design is considered to be workable. The projected plant performance is discussed and suggestions for improvements are presented.

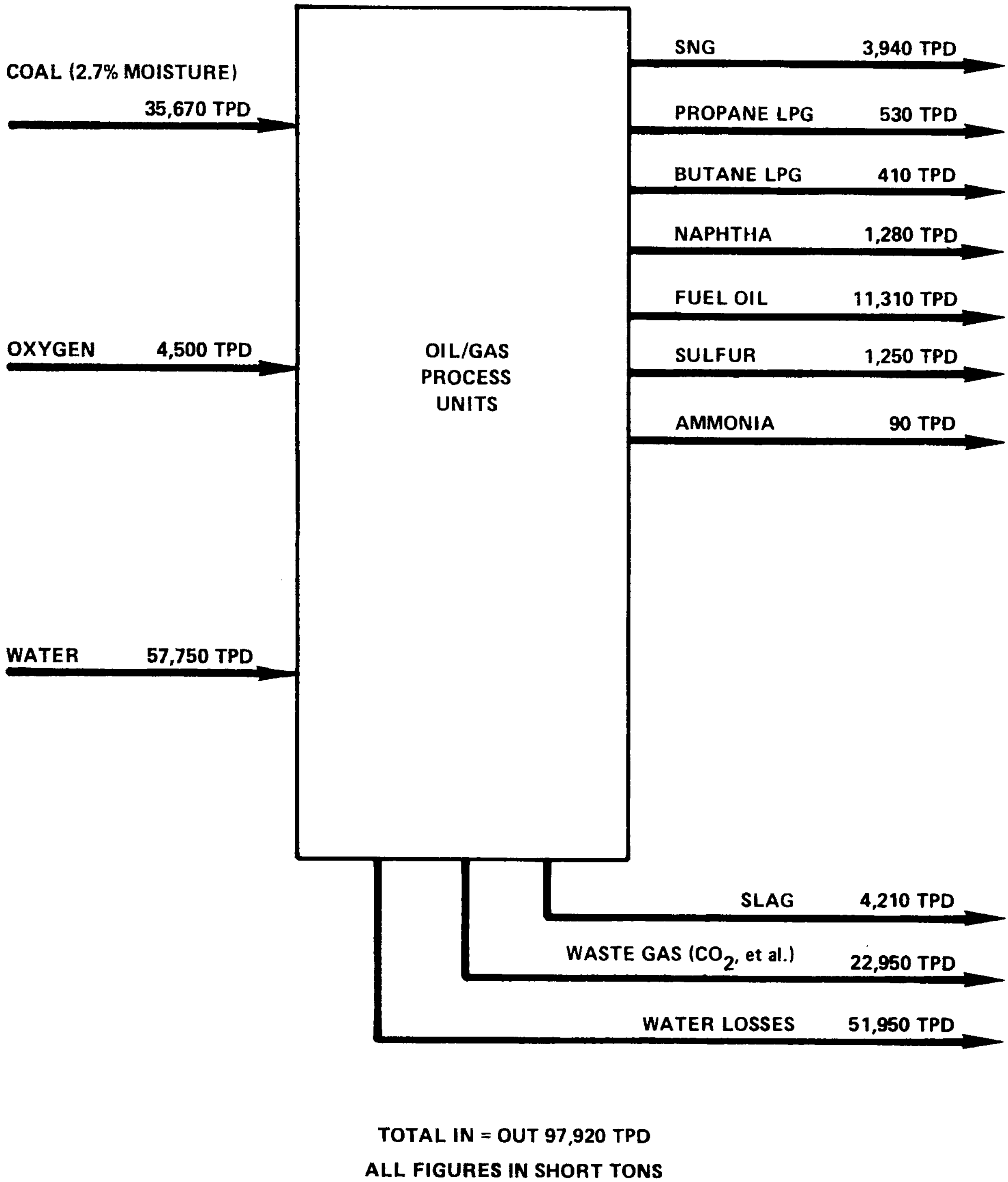

Figure 7-1 - Overall Material Balance for Complex