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A RECOMMENDED PROGRAM TO DEMONSTRATE AND PROMOTE
THE COMMERCIAL PRODUCTION OF SYNTHETIC FUELS FROM COAL

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1. INTRODUCTION AND SUMMARY

The proposed \$10 billion energy R&D program, recommended to the President on December 1, 1973, was concerned with the structuring of an orderly and well balanced effort aiming to support near term objectives while maintaining both mid-term and long-term objectives in perspective. This program recommended several large-scale projects to further develop the production of synthetic fuels from coal, including: (1) an 80 million ft³/day second generation high-Btu gas demonstration plant operational by 1980, (2) two advanced liquefaction prototype plants operational by 1978, and (3) two commercial scale pioneer plants using existing liquefaction technology for operation by 1980.

In response to more recent developments, this report provides background information in support of a proposed supplementary synthetic fuels demonstration program which is a recommended next step of an urgent national program to build productive capacity for massive quantities of liquid and gaseous fuels from coal. In view of the transportation fuels crisis and because commercial scale production of high- and low-Btu gas from coal is already being planned by several utilities, production of liquid fuels is assigned high priority.

It is the judgment of many experts that synthetic natural gas (SNG) from coal can and should be produced on a commercial scale using the existing first generation technology. Second generation processes for SNG production potentially will improve the economics (perhaps by about 20%) but these technologies are not directly required to respond to the need for increased productive capacity. It is also a general consensus that the economics of existing processes for liquefaction of coal (e.g., Bergius and Fischer-Tropsch) are so unfavorable with respect to those of the developing processes (e.g., H-Coal and SRC) that it will be preferable to build an industry based upon the new technology while conducting a co-current program to complete the necessary research and development.

The recommended synthetic fuels demonstration program would create a government-industry management system with the specific goal of providing the commercial scale development that will be required for a massive

synthetic fuels production industry in the 1980s. Specific aspects of this program include: (1) removal of governmental and other obstacles and provision of incentives to accelerate the present and follow-on commercial projects for the production of high-Btu gas from coal, (2) acceleration of the synthetic fuels pioneer program with higher government funding to compensate for the higher risk, and (3) expansion of the liquefaction prototype plant and synthetic fuels pioneer programs to increase the probability of technological success. The technology exists for extraction of the coal that would be required as a raw material for a synthetic fuels industry. Work is required, however, to minimize the impact of coal mining on the environment and to provide the necessary manpower, materials, equipment, and services for rapid expansion of the mining industry.

The following sections of this report will present an assessment of the status of the technology for liquefaction and gasification of coal and summarize the recommended synthetic fuels demonstration program.

2. SCOPE AND BASES FOR STUDY

One of the objectives of the present study has been to evaluate the technical and economic merits of the most promising near-term processes for liquefaction and gasification of coal in sufficient depth to permit a preliminary determination with respect to which of these processes should be stressed in an accelerated development program. In general, the technical and economic data used in these evaluations represented wide variations in detail, degree of substantiation, and groundrules. An attempt has been made to present a balanced technical and cost evaluation, whenever possible, through the use of only a few types of coal and common economic assumptions. In spite of this effort, however, the indicated comparisons of the processes may be in error by perhaps $\pm 30\%$.

Table 2.1 presents properties of several types of coal that are reasonably typical of those from the large coal fields in the U.S. Most of the conceptual flowsheets for this study are based upon (1) a high-volatile

Table 2-1. Compositions and Heating Value of Several Typical Coals

	Illinois, Indiana, Western Kentucky		Upper Ohio River Valley		Williston Basin		Hanna Fork Field		Powder River Basin		Four Corners	
	Clifty Creek No. 6	Illinois No. 6 River King	West Virginia Pittsburgh Seam	High-Volatile Bituminous A	Mohawk North Dakota	Lincoln County Wyoming	High-Volatile Bituminous B	Wyodak Wyoming	Sub- bituminous	Sub- bituminous	Navaho ^a FP-1	Sub- bituminous
Proximate Analysis, as received												
Moisture, wt %	12.7	11.3	2.3	36.8	22.7	29.0	12.4	16.5	13.5			
Ash, wt %	12.7	10.3	5.2	5.9	5.0	5.8	25.6	17.3	18.9			
Volatile matter, wt %	31.3	33.8	36.5	27.8	-	33.2	34.7	66.2	54.6			
Fixed carbon, wt %	45.6	44.6	56.3	29.5	-	32.1	29.0					
Ultimate Analysis, dry basis												
Carbon, wt %	66.7	70.3	80.2	64.3	76.1	66.8	54.2	60.8	57.9			
Hydrogen, wt %	4.6	4.9	5.4	4.5	5.4	5.2	4.1	4.5	4.3			
Nitrogen, wt %	1.3	0.9	1.6	1.0	1.4	0.9	1.0	1.1	1.0			
Sulfur, wt %	4.5	3.1	0.8	1.4	1.1	0.7	1.0	0.8	0.7			
Oxygen, wt %	8.4	9.2	6.6	19.4	9.5	16.3	10.5	12.1	13.5			
Ash, wt %	14.5	11.6	5.4	9.4	6.5	8.1	29.2	20.7	22.6			
Higher Heating Value												
As-received, Btu/lb	10,690	11,460	14,040	7,000	10,630	8,600	8,500	8,870	8,610			
Moisture, ash free, Btu/lb	14,330	14,620	15,180	12,220	14,700	13,180	13,710	13,400	13,310			

^a Alternative compositions used for design by Fluor Corporation, Stearns-Rogers, and Davy Powergas.

bituminous A coal from the Eastern Interior Region (Illinois, Indiana, Western Kentucky), (2) a subbituminous coal from the Powder River Basin in Wyoming, and/or (3) a subbituminous coal from the Four Corners Region in New Mexico.

Table 2.2 presents the economic parameters used in calculating the capital component of annual cost. The reference case, with a debt-equity ratio of 30/70, is relatively typical in the present petroleum and chemicals industries. The alternative case, with a 60/40 debt-equity ratio, represents a more desirable type of low-risk financing that might become possible if the synthetic fuels industry were strongly supported by the government using mechanisms such as price supports, subsidies, guaranteed loans, etc. The assumed interest rate of 8% on long-term debt is believed reasonable for large U.S. corporations with high (AA or AAA) credit ratings.

A third financing structure was assumed for estimating the cost of SNG produced by a utility company. This utility financing model (55% debt at 7.75%, 10% preferred stock at 7.75%, and 35% common stock at 14%) results in a weighted average cost of money of 10% and a fixed charge rate on capital of 16.7%.

Other economic ground rules used in this study are as follows:

1. Interest during construction was charged at 16% of the total capital investment, based on the assumption that plant construction would be financed by an intermediate term loan at an interest rate of about 8.5%.
2. Costs of as-received, mine-mouth, prepared coal at \$7.50/ton for Eastern Interior Region bituminous coals and \$3.00/ton for Western strip-mined coals.
3. All costs are expressed in terms of 1975 dollars.
4. Direct operating labor cost at \$5.00/man-hour.
5. Raw and treated water at 0.10 and 0.30 \$/1000 gal, respectively.
6. Electric power at \$0.011/kWhr.
7. 330 operating days/year.
8. 15% contingency on capital costs.

Table 2-2. Economic Parameters for Calculation
of Fixed Charges on Capital

	Reference	Low Risk
Percent Debt	30	60
Percent Common Equity	70	40
Amortization Period, years	20	20
Interest Rate on Debt, %	8.0	8.0
Earning Rate on Equity, % ^a	16.0	16.0
Fixed Charge Rate on Capital, %		
Cost of money	13.6	11.20
Depreciation ^b	1.2	1.5
Federal income tax ^c	5.5	2.3
State income tax ^d	0.3	0.2
Property taxes	2.3	2.2
Property insurance	<u>0.5</u>	0.5
Total ^e	23.4	17.9

^aAnnual after-tax rate of return on equity.

^bSinking fund depreciation.

^cFederal income tax rate 48%.

^dState income tax rate 3%.

^eThe annual fixed charges on capital are determined by multiplying the fixed charge rate by the total capital investment including interest during construction.

3. LIQUEFACTION OF COAL

The development of viable processes for the production of clean liquid fuels from coal has been assigned high priority because it has a favorable probability of technical success with acceptably low impacts on the environment and is a direct response to the goal of providing national self-sufficiency in liquid fuels. The recommended national program in coal liquefaction will continue the development of several types of liquefaction processes, develop methods for the production and use of methanol from coal as a substitute automotive fuel, investigate new types of processes, and generally provide the support in basic research, engineering development, product applications, coal mining, waste management, safety, and social and institutional endeavors that will be needed for a massive new industry.

Commercial plants for the liquefaction of coal will need to be located in the vicinity of large coal mines. The principal environmental impacts of such plants will be related to the needs for mining and conveying the coal, the disposal of solid waste (amounting to between 10% and 30% of the coal charged), and supplying the water requirements of the process. About 500 gal of water will be consumed for each ton of coal in the production of about 110 gal of liquid fuel. Current information indicates that it will be feasible to remove all significant quantities of noxious materials from plant effluents, but continued research is needed to evaluate the potential industrial hygiene and environmental problems associated with the management of gaseous, liquid, and solid wastes.

Innovative policies will be needed to solve the institutional problems that will result from the need to provide government support of the development programs while making equitable use of the proprietary experience and facilities of private industry. New legislation will be needed to solve problems associated with the substantial requirements for capital funds and manufacturing services that will be required for the buildup of a coal liquefaction industry in the 1980s. Some form of government support will be needed to reduce financial risks to an acceptable level and thereby encourage rapid growth of the industry.

Status of Technology

Present processes for the direct catalytic hydroliquefaction of coal are technological evolutions of a process invented by Bergius, which was used in 12 plants during World War II to supply approximately 85% of Germany's requirements (about 50,000 bbl/day) for aviation gasoline.¹ In the Bergius process, a slurry of coal, process-derived oil, and a small amount of iron oxide catalyst was contacted with hydrogen in a stirred reactor at a temperature of about 850°F and a pressure of 3000 (with brown coal) to 10,000 (with bituminous coal) psi. The liquid product (syncrude), which was separated from the residual char-ash-catalyst solids by distillation and carbonization, was hydrotreated further to produce high-octane gasoline. The solids, together with additional coal, were burned to provide process heat. Hydrogen required for the process was produced by fixed-bed gasification of coal with steam and oxygen. The plants were shut down after the war because comparable gasolines could be produced from imported petroleum at about 40% of the cost. Pyrolysis and Fischer-Tropsch processes for the production of liquid fuels from coal were also developed and employed during the war, but with less success than the Bergius process.

Continued research and development of processes for converting coal to clean liquids has been sponsored since the early 1960's by U.S. Government agencies and private industry. This has included: (1) development of an improved process (H-Coal) for direct catalytic hydroliquefaction; (2) construction of pilot plants to investigate multistage processes (Consol Synthetic Fuel and Solvent Refined Coal); (3) construction of a pilot plant to investigate a pyrolysis process (COED); and (4) investigative research on new processes.

Direct Catalytic Hydroliquefaction. - A significant advancement in the direct catalytic hydroliquefaction of coal, the "H-Coal" process, involving the contacting of a coal-oil slurry with a massive quantity of cobalt molybdate catalyst in a liquid fluidized bed reactor, was adapted in the early 1960's from a successful process for hydrodesulfurization of heavy petroleum fractions.² Compared with the Bergius process, such a process has promise of higher net liquid yields (up to 3 bbl/ton coal vs 1.5 for the German plants), lower operating pressure (2500 psi vs 10,000 psi for bituminous coals), and significantly lower content of sulfur in the liquid product. These technical advantages coupled with the projected escalation of the real price of low sulfur fuels from

petroleum and significant economies of scale that can be expected in commercial (~100,000 bbl/day) liquefaction plants, form the basis of a projection that such a process may be economically acceptable in the mid-1980's.

The direct catalytic H-Coal process is the most highly developed of the recent coal liquefaction processes. The process was developed by Hydrocarbon Research, Inc., (HRI) under the sponsorship of the Office of Coal Research from 1965 through 1967 under a \$2 million contract. Subsequent development to the present date, at a total cost of about \$6 million, has been sponsored by a consortium of petroleum companies. Since 1966, a process development unit having a capacity of 3 to 8 tons coal/day and bench scale units with capacity of 25 lb coal/day have been operated successfully with several types of lignites, and bituminous and subbituminous coals. This development work furnishes a substantial data base to permit the design of a large pilot (or prototype) plant and indicates a very high probability of technical success.

HRI, in cooperation with several energy companies, has proposed a joint program with the Office of Coal Research to design, construct, and operate a 300-to-700 ton coal/day prototype plant to further develop the H-Coal process for production of syncrudes and boiler fuel from coal. An extensive conceptual design has been completed for this prototype plant. This plant is considered to provide the optimal next stage (before commercialization) of scaleup of the process based upon the following considerations:

- (1) It represents a low risk and technically feasible scaleup of a factor of 100 above the present process development unit.
- (2) The plant is large enough to utilize the types of equipment (vessels, pumps, compressors, heat exchangers, filters, etc) that would be employed in a commercial plant and would require a scaleup to single trains of full-sized commercial plants by factors of only 10-to-20.
- (3) As opposed to the direct scaleup to the first train of a commercial unit (about 6000 to 10,000 tons coal/day) the prototype plant provides for (a) considerably less technological risk, (b) lower

capital investment (about \$40 million vs \$150 million), (c) faster completion (perhaps 30 months vs 50 months), and (d) greater facility for rapidly making necessary process and equipment modifications.

- (4) Such a prototype plant will be required during the operation of a commercial plant to provide a testing ground for process innovations and effects of a variety of types of coal.

Extraction-Hydrogenation Processes. - Another recent process for production of liquid fuels from coal is a modification of the Solvent Refined Coal (SRC) process.³ This process was developed originally by the Pittsburgh and Midway Co. Company for the production of a low-sulfur, ashless, pitch-like solid fuel. It is now being considered as a process for liquid fuels production through the addition of a second stage of hydrogenation. This modified process is now similar to the two-stage CSF process that was pilot-planted at Cresap, West Virginia, from 1967 to 1972. It requires dissolution of the coal in a hydrogenated oil, separation of solids from the liquids, and further catalytic hydrotreating of a portion of the liquids to achieve desulfurization and a stable liquid product. The process is primarily intended to produce heavy fuel oils suitable for industrial boilers and power plant use. A primary uncertainty with respect to this type of process relates to the as-yet undemonstrated availability of a technically and economically feasible method for separating the fine ash-char solids from the viscous and unstable liquid extract that results from the dissolution step.

Two pilot plants for development of the SRC process are planned for initial operation in 1974. Both plants, a 6-ton/day unit in Wilcoxville, Alabama (sponsored by EPRI and a consortium of private utilities) and a 50-ton/day unit in Tacoma, Washington (sponsored by OCR) are presently designed to produce a de-ashed, low-sulfur solid product (melting point 300°F) but should provide data that will permit a better assessment of the technical and economic feasibility of the required solid-liquid separations. These pilot plants are presently equipped to test precoat and rotary filters as methods for accomplishing the necessary solid-liquid separation. Plans are also being made for future tests of potentially more viable solid-liquid separation steps (e.g., hydrocyclones and agglomeration-precipitation) and catalytic hydrogenation of the product to produce liquid boiler fuels.

Old Ben Coal Company, Standard Oil of Ohio, Consolidation Coal and several other industrial sponsors have recently proposed a joint program with the EPRI and OOR to design, construct, and operate a 300 ton coal/day extraction-hydrogenation prototype (or demonstration) plant at a Sohio refinery near Toledo, Ohio.⁴ This plant has sized to provide suitable precommercial scale development of process equipment and to provide sufficiently large quantities of product to facilitate testing in large utility boilers.

Exxon Corporation has also announced plans to construct and operate a "large pilot" (or prototype) plant for converting 300 ton/day of coal to liquid products by a variation of the extraction-dissolution process at a refinery site in Baytown, Texas. Exxon currently is testing, at the process development scale of 0.5 tons of coal per day, a process that utilizes a high severity (high temperature, high hydrogen overpressure) extraction step to produce an extract that permits solid-liquid separations by distillation or carbonization.

Pyrolysis Processes. - A third main type of liquefaction process - the COED process for multistage, fluidized bed pyrolysis of coal - has been developed in a 3-ton/day pilot plant under the sponsorship of the Office of Coal Research.⁵ The pilot plant has experienced a high degree of technical success. Products from the pilot plant consist of about 20% clean gas, 20% clean liquids, and 50% high-sulfur solids. Because of their high sulfur content, the solids are useful only for the production of high-Btu or low-Btu gas, combustion in boilers with sulfur control effected by stack gas treatment, or "clean" combustion in fluid beds. This type of process cannot be fully responsive to national requirements because of the low yield of liquids. The process has potential for technical and economic viability, however, especially if an oil absorption procedure for avoiding the necessity of a solid-liquid separation proves feasible and if a companion process for economically producing high-Btu gas from the high-sulfur solids is developed. Pyrolysis processes similar to the COED process are being developed independently by TOSCO (25-ton/day pilot plant) and the Garrett Corporation (0.5-ton/day process development unit).

Other types of pyrolysis, or "hydrocarbonization," processes are being developed by U.S. Steel and Union Carbide Corporation. A potential modification of the U.S. Steel "clean coke" process⁶ would produce (1) a desulfurized char by carbonization of coal at moderately high pressures in a partial atmosphere of

hydrogen, and (2) a low-sulfur liquid fuel by direct or two-stage hydroliquefaction of a second coal feed stream. Such a type of process has the advantage that the hydrogen for the hydroliquefaction unit would be produced in the carbonizer, thus eliminating the steam-oxygen-char gasification step that is needed for the types of processes that produce only a liquid product.

The Union Carbide hydrocarbonization process has been developed through a scale of 18 tons of coal per day.⁷ This process employs hydrocarbonization in a single fluidized bed to produce up to 1.5 bbl of a desulfurized liquid fuel and the equivalent of about 1.0 bbl of high-Btu gas per ton of coal. The char is gasified to produce the hydrogen required for the hydrocarbonization and subsequent hydrotreating steps. This type of process has the marked advantage of relatively low-pressure operation.

These types of hydrocarbonization processes will not be considered in further detail in the following sections because of the rather limited data available in the open literature. The Union Carbide process, however, potentially has been sufficiently developed to permit immediate scale-up to a precommercial plant.

Fischer-Tropsch Process. - A Fischer-Tropsch process for the catalytic production of gasoline and other products from a coal-derived synthesis gas is currently being utilized in a commercial plant in South Africa.⁸ At present this type of process does not appear to be competitive with the foregoing liquefaction processes under U.S. conditions of coal costs. Although the process has the advantage of established technology, it appears to require a very large capital investment and to be relatively inefficient (~25% efficient, corresponding to a gasoline yield of about 1 bbl/ton of coal).

Other Processes. - Several other potential processes for liquefaction of coal are in an early stage of development. Processes for direct catalytic hydroliquefaction in turbulent fixed beds are being tested at the bench scale by Gulf Oil and the Bureau of Mines.⁹ Processes for hydroliquefaction in molten chloride salts are being investigated by Shell Oil and Consolidation Coal. Processes for "flash hydrogenation," rapid heating of coal in contact with hydrogen, are being investigated at the bench scale.^{10,11} Finally, engineering studies are under way to investigate the feasibility of producing methanol from coal by combining processes to produce synthesis gas from coal with already available industrial processes for production of methanol from synthetic gas.

References

1. "Report on the Petroleum and Synthetic Oil Industry of Germany," British Ministry of Fuel and Power, His Majesty's Stationery Office, London (1947).
2. K. C. Hallwig, E. S. Johnson, C. A. Johnson, S. C. Schuman, and H. H. Stotler, "Make Liquid Fuels from Coal," *Hydrocarbon Processing* 45 (5) 165-69 (May 1966).
3. "Demonstration Plant, Clean Boiler Fuels from Coal, Preliminary Design/Capital Cost Estimate," R&D Report No. 82, Interim Report No. 1, Vol. 1, prepared by the Ralph M. Parsons Company for the U.S. Department of Interior, Office of Coal Research.
4. Prospectus for Clean Fuels from Coal Demonstration Project, Old Ben Coal Corporation.
5. A. H. Strom and R. T. Eddinger, "COED Plant for Coal Conversion," *Chem. Engr. Prog.* 67 76-80, March 1971.
6. K. A. Showalter and X. S. Boodmon, "Clean Coke Process," 60th Annual Mtg. A.I.Ch.E., Philadelphia, Pa., Nov. 11-15, 1973.
7. C. W. Albright and H. G. Davis, "A Process for Dry Hydrogenation of Low Rank Coals with High Yields of Phenolics," ACS Division of Fuel Chemistry Symposium, Chicago, Sept. 13-18, 1970.
8. H. Tramm, "25-yr of F/T. Synthesis," 5th World Petrol. Congress, Section III, New York (1959), pp. 347-361.
9. P. M. Yavorsky, "Hydrodesulfurization of Coal Into Nonpolluting Fuel Oil" U.S. Bureau of Mines, Pittsburg Energy Research Center (Oct. 1972).
10. "Intermediate Coal Hydrogenation Process," in 1972 Annual Report of the Office of Coal Research, p. 75, U.S. Dept. of Interior 1973.
11. A. M. Squires, "The Coalplex: Gas, Gasoline, and Clean Electricity From Coal," 65th Annual Meeting of the A.I.Ch.E., New York, N. Y., Nov. 26-30, 1972.

3.1 Direct Catalytic Hydroliquefaction (H-Coal)

The H-Coal process has been developed by Hydrocarbon Research, Inc. (HRI), under sponsorship of the OCR and a group of oil companies. Currently, it is being studied in a process development unit at HRI's laboratory at Trenton, New Jersey. Flowsheets have been developed for the production of a syncrude product or, alternatively, a low-sulfur fuel oil product. The degree of hydrogenation of the coal determines the type of product that is produced.

3.1.1 Process Description

A flowsheet of the H-Coal process is shown in Fig. 3.1-1. Coal is crushed, ground, dried, and mixed with a coal-derived solvent at a solvent/coal weight ratio of about 1/1. After heating, the slurry is fed to an ebullating (fluidized) bed reactor, where it is contacted with a cobalt molybdate catalyst and hydrogen. Operation of the bed at a coal feed rate of about 32 lb per hour per cubic foot, a temperature up to 850°F, and a pressure up to 2700 psig yields a product consisting of gases, a range of distillate products including naphtha, middle distillates, and higher boiling fractions suitable for fuel oil plus some undissolved char and ash. This method of operation is referred to as the "syncrude" flowsheet. Increasing the coal throughput rate to about 93 lb per hour per cubic foot decreases the degree of hydrogenation, which results in a product containing a lower naphtha and middle distillate content and more fuel oil than the syncrude flowsheet.

The upward passage of the solids, liquids, and gases in the ebullated bed fluidizes the catalyst particles. However, the relative sizes of the catalyst and undissolved coal are such that a discrete level of catalyst can be maintained and the products leaving the top of the reactor are essentially free of catalyst solids. Catalyst is added to and removed from the bed continuously in order to maintain a desired level of activity. Effluents from the reactor include both a gas phase and a liquid phase. The gas phase is separated into hydrogen (for recycle) and light hydrocarbon products in an absorption system. Liquids from the reactor are separated, by distillation, into naphtha, fuel oil, slurry oil for recycle to the reactor, and a residuum fraction.

The unreacted coal and ash which end up in the residuum fraction can be removed by either of two procedures. The residuum-solids mixture can be carboni-

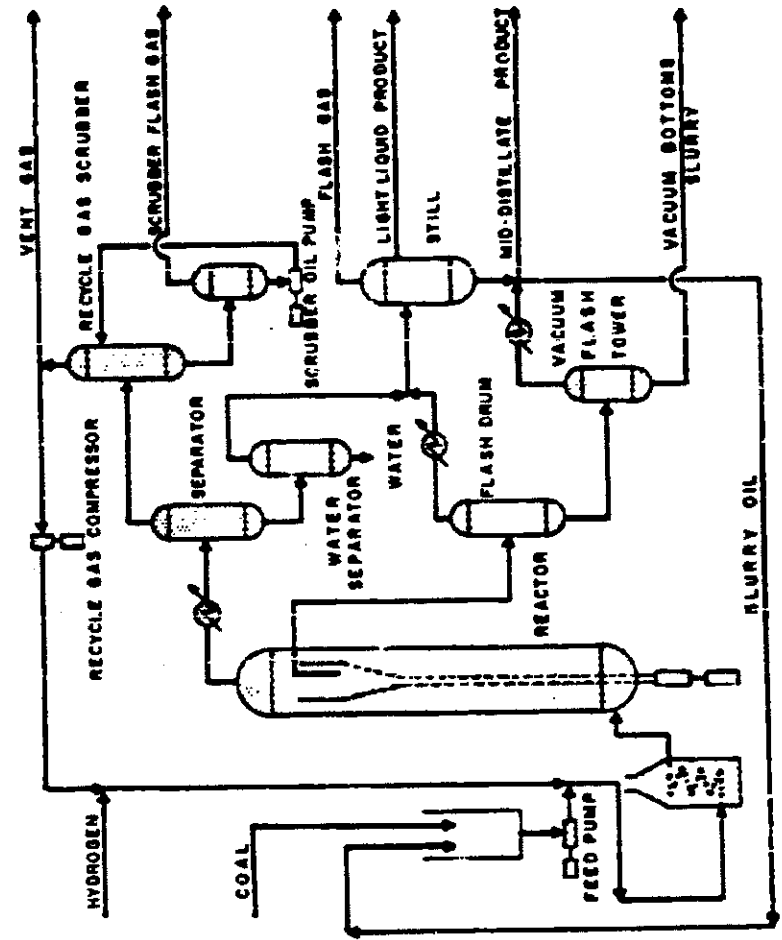


Fig. 3.1-1. H-Coal Process.

(coked) to produce distillate fuel oil plus char-ash solids. The solids can be gasified to provide plant fuel gas or hydrogen for recycle. Alternatively, the solids can be filtered from the residuum fraction and then gasified. In either case it will be necessary to scrub the gases produced from the char since they will contain high levels of H_2S .

Makeup hydrogen for the reactor can be provided from one of two internal process sources. Light hydrocarbons (C_1 to C_3) recovered from the reactor gases can be reformed with steam to generate hydrogen after the H_2S and other sulfur compounds have been removed. The second hydrogen source results from the gasification of the char-ash residue as described above. However, a pressurized gasifier with oxygen feed would be required for this step. If the char-ash mixture is used only for the production of plant fuel (hydrogen produced by reforming of the reactor off-gases), air could probably be used and the gasifier could be operated essentially at atmospheric pressure. The H-Coal process yields fuel oil products having sulfur concentrations below 0.5%.

The H-Coal process has been under development for the past eight years. However, the ebullating bed reactor has been used on a commercial scale for many years to upgrade heavy residual oils (H-Oil process). The H-Coal bench-scale units have been operated for approximately 1200 on-stream days, and the process development unit (capacity = 3-8 tons/day of coal) has been operated for 90 days since 1967. Bituminous, subbituminous, and lignite coals have been tested in the bench-scale equipment.

Several major areas of development remain before the economics of the process can be accurately ascertained. These areas include:

- a. Solids Separation. - Methods of solids separation that have been investigated include centrifugation, magnetic separation, filtration, and hydrocloning. Filtration appears to be the most promising method, but the rates achieved were quite low (30 lb/hr. ft² of filter area). For the production of synthetic crude, concentration of the solids in a vacuum tower followed by carbonization (coking) and gasification of the resulting char appears quite promising as a solids separation procedure.
- b. Char Gasification for Hydrogen Production. - Suitable methods must be developed for production of hydrogen from the undissolved coal

- solids or carbonized char in order to establish the economic feasibility of the process. Several types of gasifiers have been suggested for this service; HRI has proposed a gasifier design based upon their experience in coal gasification development.
- c. Catalyst Development. - Hydrogenation catalyst consumption is quite high when compared to normal refinery oil hydrogenation experience. Development of procedures for reducing the catalyst poisoning or regeneration of spent catalyst would substantially enhance the economic suitability of the H-Coal process.
 - d. Waste Treatment Facilities. - Demonstration of efficient waste treatment procedures for the gaseous and aqueous wastes from the H-Coal process will be a necessary part of the development program.
 - e. Materials of Construction. - Long-term testing of the materials of construction with respect to corrosion, erosion, and hydrogen embrittlement will be an essential part of the development program.

Based upon the development studies to date, the H-Coal process appears to be technically feasible and ready for scale-up to a large pilot plant or prototype plant size. The yields of synthetic coal liquids (2-3 bbl/ton of dry coal) and process efficiency are acceptable. The projected economic analysis appears favorable.

3.1.2 Characteristics of Conceptual Prototype, Demonstration, and Commercial Plants

Hydrocarbon Research, Inc. has prepared conceptual designs for a series of facilities including: (1) a 250-700-ton coal/day prototype plant for producing syncrude or fuel oil products, (2) a 4000-ton coal/day demonstration plant for producing syncrude, (3) a 26,000-ton coal/day commercial plant for producing syncrude from Illinois No. 6 coal, and (4) a 30,000-ton coal/day commercial plant for the production of syncrude from Wyoming Wyodak subbituminous coal. Material and energy balance data for representative plant designs based upon the HRI data are shown in Table 3.1-1. The yields of liquid from coal increase in the larger scale plants due to the increased efficiency and the utilization of by-product char in the production of hydrogen (for cases 2, 3, and 4). The lower yield for the Wyodak plant is due in part to the high moisture content of the Western coal. The yields shown here for the

Table 3.1-1. Estimated Raw Materials, Utilities, and Products of Conceptual H-Coal Prototype, Demonstration, and Commercial Plants
330 Operating Days per Year

Case No.	1	2	3	4
Plant Type	Prototype	Demonstration	Commercial	Commercial
Coal Type	ILL. No. 6	ILL. No. 6	ILL. No. 6	Wyodak
Reference	2	3, 4	1	1
Raw Materials and Utilities				
As received coal, t/d	1044	5221	35,274	50,466
Dry coal, t/d	940	4699	31,747	35,832
Electricity, KW	7450	39,000	257,000	333,000
Water, GPM	485	3300	14,100	15,500
Energy input, 10 ⁹ Btu/d ^a	23.2	115.7	781.7	909.2
Products and Effluents				
Type of product	Fuel oil + naphtha	Fuel oil + naphtha	Synchrude	Synchrude
High Btu fuel gas, 10 ⁹ Btu/d	-	-	109.9	131.5
Total liquid products, bbl/d	2258	12,974	82,508	79,120
Total products, bbl ^b /ton dry coal feed	2.40	2.76	3.15	2.79
Naphtha, vol %	17.5	36.9	42.2	64.0
Middle distillate, vol %	82.5	63.1	41.5	34.0
Heavy gas oil, vol %	-	-	16.3	2.0
Gravity, °API	~15	~23	25.2	39.3
Sulfur, wt %	<0.2	<0.2	<0.2	<0.2
Heating value, 10 ⁹ Btu/d	13.8	76.11	469.9	443.4
Sulfur, t/d	17.7	195	1315	295
Ammonia, t/d	-	30	251	87.3
Waste water, t/d	800	6000	29,400	28,000
Solid wastes (ash, catalyst), t/d	119.4	700	4100	4600
Process Efficiency, %	60	66	74	63

^a All heats of combustion based upon higher heating value.

^b Fuel gas expressed as equivalent barrels of fuel oil.

H-Coal process equal or surpass those for other proposed liquefaction processes.

The efficiencies of the H-Coal plants vary between 60 and 74% (efficiencies are expressed as the ratio of product heating value to total coal feed heating value). Again, the lower efficiencies are due to operation in a smaller plant or consumption of coal with a high moisture content. The efficiency results of 60-74% compare favorably with other coal liquefaction plants and surpass those estimated for substitute natural gas plants.

The estimated capital investments required for the conceptual H-Coal plants are tabulated in Table 3.1-2. These investments are for grass-roots plants and include all the required facilities except an electrical generation plant. It was assumed that power would be available from a nearby more efficient central station. A 15% contingency has been added to the HRI estimates to cover uncertainties in the costs of items such as filters and gas generators. Working capital has been estimated at 10% of fixed capital investment. The data shown in Table 3.1-2 are for 1973 and have not been corrected for escalation to the actual purchase or installation dates. The cost of the coal mine is not included.

Annual operating costs are presented in Table 3.1-3. These estimates were based upon HRI information corrected for the economic ground rules of this study. The annual revenues required to match the annual operating costs were used to estimate the unit production costs for the liquid products. The high-Btu gas produced in the commercial plants was costed at the same price as the liquid products per unit of heating value.

Manpower and critical resource estimates for the H-Coal plants are shown in Table 3.1-4. Design and construction manpower requirements were scaled up from estimates prepared by The Ralph M. Parsons Company for their study of a 10,000-ton/day demonstration plant for clean boiler fuels.⁵ Operations personnel were estimated from direct and contract maintenance costs developed by HRI and personnel requirements listed by HRI for the prototype plant.

Table 3.1-2. Estimated Capital Investment (1973 dollars) for Conceptual H-Coal Prototype, Demonstration, and Commercial Plants

Case No.	1	2	3	4
Plant Type	Prototype	Demonstration	Commercial	Commercial
Coal Type	Ill.No.6	Ill.No. 6	Ill. No. 6	Wyodak
Reference	2	3,4	1	1
<u>Capital Costs, \$ millions</u>				
On-Site				
Coal Preparation	1.78	5.5		
Hydrogenation	11.27	36.0		
Reforming and H ₂				
Compression	2.73	21.0		
Coking	1.60	6.5		
Coal-Char Gasification	13.90	32.0		
Gas Clean-up	0.33	3.0		
Waste Water Treatment	0.78	1.6		
Sulfur Manufacture	0.92	3.4		
	<u>33.31</u>	<u>109.0</u>	<u>366.9^b</u>	<u>449.3^b</u>
Off-Site				
Utilities	1.50	4.7		
Buildings	1.31	1.7		
Tankage	0.56	4.1		
Electrical Substations	1.57	5.0		
Land and Land Rights				
Site Preparation				
Miscellaneous	0.75	2.1		
	<u>5.69</u>	<u>17.6</u>	<u>66.0^b</u>	<u>65.5^b</u>
Engineering	^a	^a	^a	^a
Taxes	1.35	4.39	15.10	17.93
Fees	^a	^a	^a	^a
Catalysts and Chemicals	0.10	0.70	4.9	4.8
Contingency at 15%	5.87	19.10	65.7	77.9
Start-up Costs	1.96	5.93	28.6	23.1
Interest During Construction	7.72	25.08	87.6	102.2
Initial Working Capital	<u>4.50</u>	<u>14.6</u>	<u>50.4</u>	<u>59.8</u>
Total Capital Investment ^c	<u>60.5</u>	<u>196.4</u>	<u>685.2</u>	<u>800.5</u>

^aEngineering and fees included in on-sites and off-sites.

^bBreakdown not available.

^cCost of coal mine is not included.

Table 3.1-3. Estimated Annual Operating Costs and Revenue for Conceptual H-Coal Prototype, Demonstration, and Commercial Plants
330 Operating Days per Year

Case No. Plant Type Coal Type Reference	1 Prototype Ill. No. 6 2	2 Demonstration Ill. No. 6 3, 4	3 Commercial Ill. No. C 1	4 Commercial Ill. No. C 1
Annual Costs (\$ millions)				
Direct labor payroll	3.82	2.10	4.93	5.25
Support labor payroll		2.10	2.18	2.20
Overhead and burden		7.07	8.20	10.20
Contracted services	1.07	12.92	6.30	7.62
Maintenance materials & equipment	1.42	3.40	87.30	49.96
As-received coal ^b	2.58	1.90	22.35	29.04
Electricity at 0.011/kWhr	0.65	0.16	11.61	11.43
Catalysts and chemicals	0.22	0.02	-	-
Treated water	0.02	14.16	160.34	187.32
Capital charges at 23.4%	14.16	75.61	303.21	303.02
Total	23.94	75.61	303.21	303.02
Annual Revenue (\$ millions)				
High-Btu Gas	-	75.36	57.08	69.15
Syncrude	23.94	-	244.06	233.15
Sulfur at zero value	-	-	0	0
Ammonia at \$25/t	-	0.25	2.07	0.72
Unit Production Costs, \$/10 ⁶ Btu				
High-Btu Gas, \$/10 ⁶ Btu	5.27 ^a	3.04	1.57	1.59
Liquid products, \$/10 ⁶ Btu	32.13	17.85	1.57	1.59
Liquid products, \$/bbl			8.94	8.93

^a Product is sold as boiler fuel oil instead of Syncrude.

^b Illinois No. 6 coal @ \$7.50/ton, Wyodak coal @ \$3.00/ton.

Table 3.1-4. Manpower Requirements for Design, Construction, and Operation and Requirements for Steel in Construction

Case No.	1	2	3	4
Plant Type	Prototype	Demonstration	Commercial	Commercial
Coal Type Reference	Illinois No. 6 2	Illinois No. 6 3, 4	Illinois No. 6 1	Wyodak 1
Design Engineering Manpower, man-hours				
Direct: Monthly, weekly, and hourly	200,000	660,000	2,250,000	2,680,000
Support: Monthly, weekly, and hourly				
Total				
Construction Manpower, man-hours				
Direct: Monthly, weekly, and hourly	810,000	2,650,000	9,000,000	10,700,000
Support: Monthly, weekly, and hourly				
Total				
Operations, number of employees				
Direct: Monthly	28	160	347	369
Weekly and hourly	76			
Support: Monthly	19	142	307	326
Weekly and hourly	73			
Contract: Monthly	21	140	325	405
Weekly and hourly	62			
Total	279	442	979	1,100
Requirements for Steel, tons				
Equipment	8,200	26,500	90,000	107,000
Structural	2,000	6,600	22,000	27,000
Total	10,200	33,100	112,000	134,000

3.1.3 References

1. Hydrocarbon Research, Inc., The H-Coal Process, Laboratory Report No. P-73-505-P (June 15, 1973).
2. Hydrocarbon Research, Inc., Prototype Plant Program: 250-700 Tons per Day with HRI Coal Gasification, Laboratory Report No. P-73-505-SG (June 15, 1973).
3. Hydrocarbon Research, Inc., Prototype Plant Program: 4000 Tons per Day with HRI Coal Gasification and Refinery Gas, Laboratory Report No. P-73-505-LG (June 15, 1973).
4. Hydrocarbon Research, Inc., "Synthetic Fuels from Coal," unpublished report (1973).
5. "Demonstration Plant, Clean Boiler Fuels from Coal, Preliminary Design/Capital Cost Estimate," R&D Report No. 82, Interim Report No. 1, Vol. 1, prepared by The Ralph M. Parsons Company for the U.S. Department of the Interior, Office of Coal Research.

3.2 Extraction - Hydrogenation

Process development effort has been directed toward the production of clean solid and liquid boiler fuels from coal by a combination of non-catalytic dissolution (or "extraction") in a hydrogenated process derived oil and a second stage of catalytic hydrogenation. Two proposed projects for exploitation of this type of process are discussed in this section. These are the Clean Fuels from Coal project (CFC) proposed by the Old Ben Coal Corporation¹ and the Clean Boiler Fuels from coal project (CBF) that has been proposed by the Ralph M. Parsons Company.²

3.2.1 Process Description, Evaluation, and R&D Needs

The projects discussed in this section are similar in concept although they differ in specific details. In general the process can be subdivided into the following major sections.

- Coal Preparation
- Coal Liquefaction
- Solid-Liquid Separation
- Liquid Product Distillation
- Product Hydrogenation
- Gasification
- Gas Treatment
- Sulfur Recovery
- Oxygen Production
- Hydrogen Production
- Waste Treatment

As an illustration of the arrangement of these sections, a schematic flow diagram is shown in Fig. 3.2-1 for the CBF project. The coal is dried, ground and mixed with a recycle solvent. The coal-solvent slurry is fed to a reactor where liquefaction is accomplished by direct hydrogenation and by hydrogen transfer by solvent donation. Variations in product distribution can be achieved by adjusting the severity of conditions to achieve greater or lesser hydrogen input. The operating temperature is in the range of 750°F to 850°F and the operating pressure may vary from 1000 psi to 2000 psi.

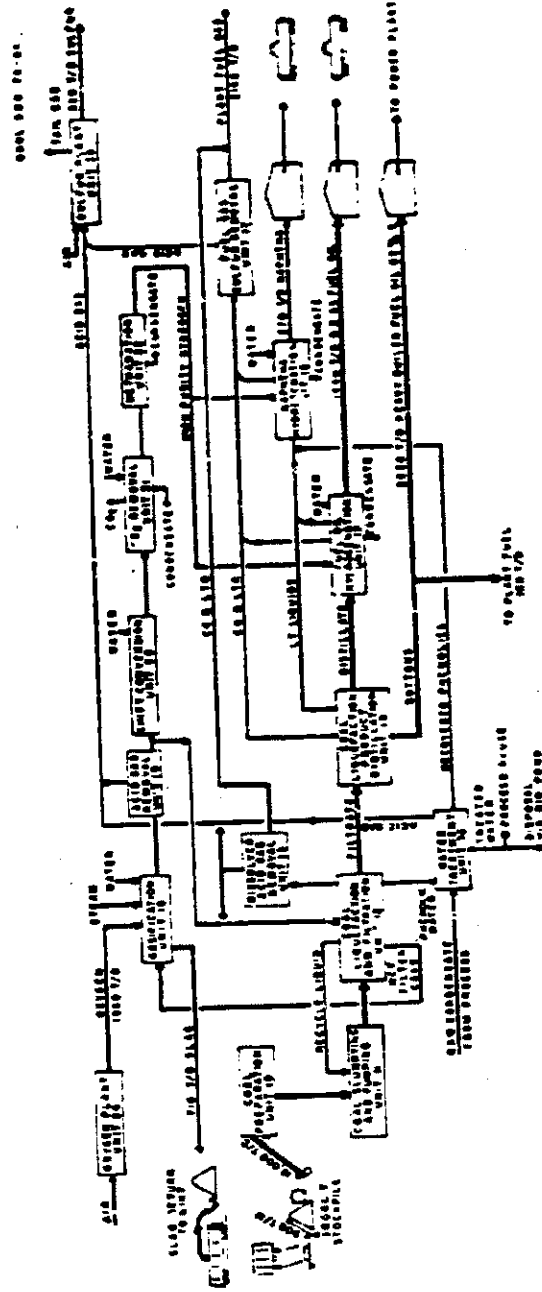


Fig. 3.2-1. Schematic Flow Diagram of Clean Boiler Fuels from Coal Process for Demonstration Plant.

Two methods have been proposed for the solid-liquid separation. The CFC project proposes the use of hydroclones to accomplish this separation. There is some concern that incomplete separation may result in sufficient carryover of solids to cause plugging problems in subsequent processing equipment. The CBF project utilizes rotary precoat filters for this separation based on FMC Corporation's experience. There is some concern about the effect of operating parameters on filtration rates and the ability to scale experience on small units to large equipment sizes. An alternate to filtration may be a stripping procedure; however, no details on this method are currently available.

The CBF project utilizes two stages of hydrogenation. In addition to the liquefaction step, the naphtha and fuel oil streams are further hydrogenated to improve product quality.* The CFC project utilizes hydrocracking of the extract to increase production where liquid fuels are the desired product.

The CBF project uses a slagging, suspension-type gasifier that is a modification of the Bi-Gas unit presently under development by the Bituminous Coal Institute. Several problem areas that may exist with this type unit are slurry feeding, overhead dust removal, slag removal, and refractory degradation. It appears that there is sufficient industrial experience with these problem areas to permit the design of the gasifier units. The better-developed Lurgi gasifiers would not be suitable for the CBF process without extensive additional equipment. Koppers-Totzek gasifiers could be used but this would require a large number of units.

Research and development effort is needed in several areas to facilitate commercialization of the extraction-hydrogenation process. The problems associated with solid-liquid separation techniques have been mentioned previously. Experience of commercial size equipment is needed to determine the extent of the problems and to develop any solutions that may be required. Operational experience with the slagging, suspension-type gasifier is needed to demonstrate that it can be successfully used with the CBF process. Product evaluation is needed to determine the severity of hydrogenation required and the subsequent refining needs before ultimate

* If desired, hydrotreating of the naphtha and fuel oil products could also be done at a petroleum refinery rather than at the CBF plant.

disposition. Evaluation of waste treatment requirements and materials of construction will be required in the development program.

Based on current technology, it appears feasible to proceed with a prototype plant. Alternative processing options appear to exist for the major problem areas, and the major concern is to establish where these alternatives will be required for viable commercial plants. In view of these considerations and the desire to expedite commercial development of the process, it appears desirable to proceed with a prototype plant (similar in size to the Old Ben Coal Company proposal) rather than a very large demonstration plant. It is also recommended that the Wilsonville and Tacoma pilot plants (due for initial operation in 1974) be modified as necessary to permit development of alternative solid-liquid separation procedures, methods for performing the second stage of catalytic hydrogenation, and methods for hydrogen generation by gasification of the separated solids.

3.2.2 Characteristics of Conceptual Prototype, Demonstration, and Commercial Plants

Table 3.2-1 gives the estimated raw materials, utilities, and products for several cases. Shown are conceptual design numbers for a 900 T/D CFC plant producing primarily a low-sulfur distillate and for a similar plant producing primarily a clean solid fuel. Also shown are conceptual design numbers for a 10,000 T/D CBF demonstration plant and a commercial CBF plant. The capital investments required for these cases are given in Table 3.2-2. The capital investment numbers for the commercial size CBF plant were obtained by scaling costs established by Parsons² for the 10,000 T/D plant. The scale factors (exponent for the ratio of plant capacities) were estimated for each equipment category. The resulting average scale factor was 0.75. Estimated annual operating costs and revenues for the conceptual plants are given in Table 3.2-3. Table 3.2-4 lists the estimated manpower and steel requirements for these plants.

Table 3.2-1. Estimated Raw Materials, Utilities, and Products for Conceptual Extraction - Hydrogenation Prototype, Demonstration, and Commercial Plants

Case No.	1	2	3	4
Plant Type	Prototype	Prototype	Demonstration	Commercial
Process	CFC - Low Sulfur Distillate Illinois No. 6	CFC - Clean Solid Fuel Illinois No. 6	CBF Illinois No. 6	CBF Illinois No. 6
Coal Type Reference	1	1	2	
Raw Materials and Utilities				
As received coal, tons/day	1144	985	10,940	43,760
Dry coal, tons/day	1046	900	10,000	40,000
Catalyst and chemicals, tons/day	-	9400	32	128
Electricity, kW	10,900	-	(76,000) ^a	(304,000) ^a
Water, gpm	27.1	-	3,626	14,500
Energy Input, 10 ⁹ Btu/day	-	23.3	250.8	1003.2
Products and Effluents				
Fuel gas, 10 ⁹ Btu/day	6.2	-	25,380	101,520
Total liquid product, bbl/day	2100	605	7.9	7.9
Naphtha, vol. %	-	100	33.3	33.3
Middle distillate, vol. %	-	-	58.8	58.8
Heavy gas oil, vol. %	-	-	-	-
Solid fuel, tons/day	17	515	320	1280
Sulfur, tons/day	-	17	-	-
Ammonia, tons/day	-	-	19,430	77,720
Waste gases, tons/day	-	-	6,390	25,560
Waste water, tons/day	-	-	1,150	4,600
Solid wastes, tons/day	150	103	-	-
Sulfur content:				
Fuel gas	nil	-	-	-
Naphtha	-	-	1 ppm	1 ppm
Middle distillate	0.3 wt %	0.3 wt %	0.2 wt %	0.2 wt %
Heavy gas oil	-	-	0.5 wt %	0.5 wt %
Solid fuel	-	0.9 wt %	-	-
Process efficiency ^b	65%	84%	63.5%	63.5%
Total Output, 10 ⁹ Btu/day	17.7	19.5	160.1	640.4

^aGenerated internally with low-Btu gas turbines.

Extraction - Hydrogenation Prototype, Demonstration, and Commercial Plants

Case No. Plant Type Process	1 Prototype CFC - Low Sulfur Distillate Illinois No. 6 1	2 Prototype CFC - Clean Solid Fuel Illinois No. 6 1	3 Demonstration CBF Illinois No. 6 2	4 Commercial CBF Illinois No. 6
Coal Preparation			6.645	
Coal Slurry and Pumping			0.480	
Coal Liquefaction and Filtration			61.311	
Dissolver Acid Gas Removal			15.000	
Product Distillation			1.349	
Fuel Oil Hydrogenation			13.446	
Naphtha Hydrogenation			1.341	
Fuel Gas Sulfur Removal			1.479	
Gasification			13.235	
Acid Gas Removal			4.293	
Shift Conversion			4.225	
CO ₂ Removal			1.864	
Methanation			0.309	
Sulfur Plant			6.134	
Oxygen Plant			37.590	
Instrument and Plant Air			0.395	
Raw Water Treatment			5.700	
Waste Water Treatment			0.873	
Product Storage			5.888	
Slag Removal System			0.069	
Steam Generation			3.474	
General Facilities			1.900	
Total Equipment Cost			66.368	187.0
Total Construction Cost		42.2 ^a	168.3	474.4
Home Office Engineering			23.9	67.3
Contingency (15%)		6.3	28.8	81.3
Start-up Costs		1.3	9.7	34.9
Interest During Construction		8.0	36.9	105.3
Working Capital		3.5	21.0	69.0
Total Capital Investment	61.5	61.3	288.6	832.2

^a Breakdown not available.

TABLE 3.4-1. ESTIMATED ANNUAL OPERATING COST AND REVENUE FOR CONCEPTUAL
EXTRACTION - HYDROGENATION PROTOTYPE, DEMONSTRATION, AND COMMERCIAL PLANTS -
330 OPERATING DAYS PER YEAR

Case No.	1	2	3	4
Plant Type	Prototype	Prototype	Demonstration	Commercial
Process	CFC - Low	CFC - Clean	CBF	CBF
Coal Type	Sulfur Distillate	Solid Fuel	Illinois No. 6	Illinois No. 6
Reference	1	1	2	6
Annual Costs, \$10 ⁶				
Direct Labor	0.73	0.73	2.18	4.16
Supervision @ 15%	0.11	0.11	0.33	0.62
Overhead @ 60% of Labor & Supv.	0.50	0.50	1.51	2.87
Maintenance @ 4% Fixed Capital Investment	1.94	1.94	8.84	24.92
As Received Coal @ \$7.50/ton	2.83	2.43	27.08	108.31
Electricity @ \$0.011/kwhr	0.95	0.82	6.62	26.48
Catalyst and Chemicals	0.15	0.15	1.65	6.60
Raw Water @ \$0.10/1000 gal	0.02	0.02	0.17	0.68
Treated Water @ \$0.30/1000 gal	-	-	-	-
Net Operating Cost	7.23	6.70	48.38	174.64
Fixed Capital Charge @ 23.4% of Total Capital Investment	14.39	14.34	67.53	194.73
Total Annual Cost	21.62	21.04	115.91	369.37
Annual Revenue, \$10 ⁶				
Fuel Gas	7.57	-	7.38	23.53
Naphtha	-	-	37.58	119.74
Fuel Oil	14.05	3.56	70.95	226.10
Heavy Liquid Boiler Fuel	-	-	-	-
Solvent Refined Coal	-	17.48	-	-
Total Annual Revenue	21.62	21.04	115.91	369.37
Unit Production Cost				
\$/10 ⁶ Btu	3.701	3.270	2.242	1.786
\$/bbl, fuel oil equivalent	22.95	20.27	13.90	11.07
@ 6.2 x 10 ⁶ Btu/bbl				

Operation and Requirements for Steel in Construction

Case No.	1	2	3	4
Plant Type	Prototype	Prototype	Demonstration	Commercial
Process	CFC - Low Sulfur Distillate Illinois No. 6	CFC - Clean Solid Fuel Illinois No. 6	CBF Illinois No. 6	CBF Illinois No. 6
Coal Type Reference	1	1	2	
Design Engineering Manpower, Man-hours	190,000	190,000	900,000	2,500,000
Direct, Monthly				
Direct, Weekly & Hourly				
Support, Monthly				
Support, Weekly & Hourly				
Total				
Construction Manpower, Man-hours	760,000	760,000	3,600,000	10,100,000
Direct, Monthly				
Direct, Weekly & Hourly				
Support, Monthly				
Support, Weekly & Hourly				
Total				
Operations, Number of People	6	6	16	30
Direct, Monthly	70	70	210	400
Direct, Weekly & Hourly				
Support, Monthly				
Support, Weekly & Hourly				
Contract Maintenance, Monthly	40	40	180	500
Contract Maintenance, Weekly & Hourly				
Total				
Requirements for Steel, Tons				
Equipment	8,800	8,800	40,000	120,000
Structural	2,200	2,200	10,000	30,000
Total	11,000	11,000	50,000	150,000

3.2.3 References

1. "Prospectus for Clean Fuels from Coal Demonstration Project,"
Cid Ben Coal Corp.
2. "Demonstration Plant, Clean Boiler Fuels from Coal, Preliminary
Design/Capital Cost Estimate," R&D Report No. 82, Interim Report
No. 1, Vol. 1, prepared by The Ralph M. Parsons Company for the U.S.
Department of Interior, Office of Coal Research.

3.3 Fischer-Tropsch Process for Gasoline

Since the description of the hydrogenation of CO to form methane in 1902 by Sabatier and Senderens and subsequent improvements in the early 1920s by Fischer and Tropsch of the Kaiser Wilhelm Institute for Coal Research, an enormous amount of research and development has been expended on commercialization of various versions of what came to be known as the Fischer-Tropsch process. The economies continue to be elusive, yet the steady interest in the process may be readily explained by its versatility and adaptability to varying economic needs and conditions of raw material supply.

Initial interest in Germany was expressed by the Ruhr Coal Syndicate in the early 1930s for the purpose of developing a market for surplus coke. Subsequently the emphasis changed to the manufacture of gasoline and synthetic oil for the war effort. However, direct hydrogenation of coal provided the major portion of the need with Fischer-Tropsch synthesis supplying much smaller amounts of gasoline and oil. Up to 1939, nine Fischer-Tropsch plants were erected with a total rated capacity of 740,000 metric tons/year and supplied 8% of Germany's home production.¹ Of these, four are evidently still in operation (three in East Germany) supplying coal chemicals; the remainder were destroyed in wartime bombings.

Interest in the U.S. in the 1930s centered around finding a use for the surplus natural gas which was a by-product of petroleum production.² Toward this objective Carthage Hydrocol built a Fischer-Tropsch plant at Brownsville, Texas, which was completed in 1950, with a rated capacity of 360,000 tons/year of gasoline and fuel oil. The process chosen was one which had been under development at the M. W. Kellogg Company since 1937 and differed from the German process in the means of effecting the contact between synthesis gas and solid catalyst. The Kellogg process suspends the catalyst particles in a gas stream which simplifies heat removal whereas the Arge reactor (developed by Lurgi) employs a fixed catalyst bed. The Brownsville plant was initially plagued by technical difficulties and changed ownership twice. By 1957, the technical difficulties (described in ref. 3) were overcome and successful operation was claimed. However, by that time the

price of natural gas had risen to the point where the process could not be operated economically and the plant was shut down.

The Synthetic Liquid Fuels Act of 1944 authorized the Bureau of Mines to conduct research and development on the production of oil from coal and oil shale, including the building of demonstration plants. An \$87.5 million expenditure was authorized over an 11-year period. The program included coal hydrogenation, coal gasification, purification of synthesis gas, Fischer-Tropsch synthesis, and mining and retorting of oil shale. The work conducted during this period on basic Fischer-Tropsch chemistry and catalyst selection is described by Storch.¹⁰ Process development work related primarily to the "oil circulation process." A demonstration plant based on this process was completed in 1950 at Louisiana, Mo. One hundred days' operation during the period 1951-1953 produced 40,000 gal of oil. The system was not considered to show promise. The Synthetic Liquid Fuels Program ended in 1955, and Fischer-Tropsch synthesis work has been carried out at the Bureau of Mines since then at a much lower rate of expenditure.¹¹

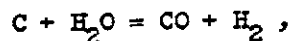
Cost estimates made in 1952 by a committee of the National Petroleum Council, using data supplied by the Bureau of Mines, showed a unit production cost of 29.1 ¢/gal for Fischer-Tropsch gasoline produced at a scale of 3800 tons/day coal utilization. At 1973 levels, this amounts to about 75¢/gal (31.45 \$/bbl) assuming an inflation rate according to Nelson's Inflation Index.

The unique situation in the Republic of South Africa set the stage for the largest Fischer-Tropsch synthesis plant presently in existence, the so-called SASOL plant located 50 miles south of Johannesburg. South Africa has no commercially exploitable petroleum and only limited supplies of high-quality bituminous coal. The large quantities of high-ash, sub-bituminous coal which were available were evidently thought unfit for direct hydrogenation to liquids by processes available at that time, i.e., the late 1940s. Thus a process based on gasification of the available high-ash coals to obtain synthesis gas and Fischer-Tropsch synthesis to yield useful liquid products was decided upon. The project was completed in 1955 and now supplies about 20% of South Africa's gasoline. Approximately 4900 bbl/day of useful product are produced, of which 76% is gasoline. The yield is 1.1 bbl liquid/ton m.a.f. coal.⁴

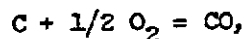
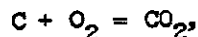
A summary⁵ of commercial plants for the production of liquid fuels via Fischer-Tropsch synthesis is presented in Table 3.3-1.

3.3.1 Process Description and Evaluation

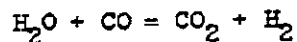
Process Chemistry. - The initial step in the process is the production of clean synthesis gas from coal with the desired H₂/CO ratio. This is identical with the initial step for methanol production which effectively is a variant of the Fischer-Tropsch synthesis reaction. The principal gasification reaction is



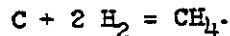
which is highly endothermic, the energy being supplied by the reactions



and also the oxidation of volatile matter in the coal. If steam is present in excess, as it is in the Lurgi gasifier in order to keep the ash below its softening point, some quantity of toxic phenols is produced as well as additional H₂ via the water gas shift



at temperatures above ~1500°F. If the gasifier pressure is significantly above 1 atm, direct methanation occurs in low-temperature zones (<1000°F) of the gasification reactor:



The Lurgi gasifier makes about 9% methane⁶, which is completely inert to synthesis and must be converted by reforming.

The Fischer-Tropsch hydrocarbon synthesis reaction may be thought of as manufacturing CH₂ groups via

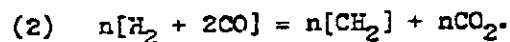
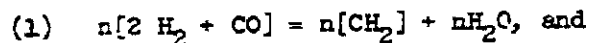


Table 3.3-1. Commercial Plants for the Production of Synthetic Liquid Fuels via the Fischer-Tropsch Process

Country	Name of Company	Startup, year	Starting material	Normal-pressure synthesis	Medium-pressure synthesis			Catalyst	Design capacity, 1000 ton/yr.
					Double tube	Fixed bed	Fluid bed		
Germany	Brabag	1936	lignite	x			x	210	
Germany	Gewerkschaft Viktor	1936	coke, coke gas	x			x	50	
Germany	Ruhrbenzin A.G.	1936 1938	coke	x	x		x	30 42*	
Germany	Gewerkschaft Rheinpreussen	1936	coke, coke gas	x			x	75	
Germany	Krupp-Treibstoffwerke	1937, 1939 1939, 1952	coke	x	x		x	47 13	
Germany	Wintershall A.G.	1937, 1939	lignite	x			x	80	
Germany	Essener Steinkohle Chemischwerke	1937, 1938	coke, coke gas	x			x	50	
France	Harnes	1937	coke	x			x	30	
	Societe Kuhlmann	1952			x			2	
Germany	Hoesch-Treibstoffwerke	1939	coke	x	x		x	60	
Germany	Schaffgotsch-Werke	1939	coke, coke gas	x	x		x	60	
Japan	Mike	1940		x			x	40	
Japan	Takikawa	1942, 1943	coal	x			x	100	
Japan	Rumoi	1939		x			x	50	
Japan	Anagasaki	1939		x			x	70	
Manchuria	Fushin	1939	coal	x			x	50	
U.S.	Hydrocol Co.	1950	natural gas			x		360	
South Africa	Sasol	1955	coal			x		185	
South Africa	Sasol	1955	coal		x			69	

It should be emphasized that the above reactions are only two of many involving the reactive compounds H_2 and CO . Many of this class of reactions form oxygenated compounds, such as methanol. All are symbolic representations of net products since the true chemical steps involved in Fischer-Tropsch synthesis remain obscure.⁷

Whether synthesis occurs via reaction (1) or (2) or both is determined by the H_2/CO molar ratio in the feed. For an H_2/CO ratio greater than 2, synthesis occurs solely via reaction (1); if the H_2/CO ratio is less than 0.5, synthesis is entirely by reaction (2).⁸ If the feed is of intermediate composition, both reactions take place, and both CO_2 and H_2O are produced.

Process Description. - The nature of the products depends on the catalyst and conditions in the reactor. Figure 3.3-1 (from ref. 7) indicates the range of possible catalysts and products.

The Kellogg Synthesis Process (also termed Synthol) appears to offer the highest gasoline yield relative to by-product production. The Kellogg Synthesis Process is carried out at $630^\circ F$ and 295 psig on sintered iron catalyst entrained in the gas stream. A typical yield for a 25,500-bbl/day plant is given in Table 3.3-2.⁹ As noted gasoline (including about 5% ethanol) comprises 73% of the product.

A block diagram for manufacture of gasoline by Fischer-Tropsch Synthesis using the Kellogg Process is shown in Fig. 3.3-2. This is closely related to the SASOL flowsheet; however, it is important to call attention to several major differences. First, process steam and power are produced using desulfurized fuel gas made on site to comply with air pollution regulations, whereas SASOL simply burns unprocessed coal. Hence, by comparison on an equal basis with SASOL, the flowsheet in Fig. 3.3-2 would show a relatively higher gasifier capital cost. Additionally, SASOL utilizes a fixed bed synthesis (Arge synthesis) to manufacture approximately 20% of the total product, principally the heavier hydrocarbon fractions. A third point of departure is the absence in the Fig. 3.3-2 flowsheet of the degree of downstream chemical processing of by-product materials which exists at SASOL, e.g., SASOL manufactures ammonium nitrate using a quantity of ammonia beyond that normally produced in the gasifiers.

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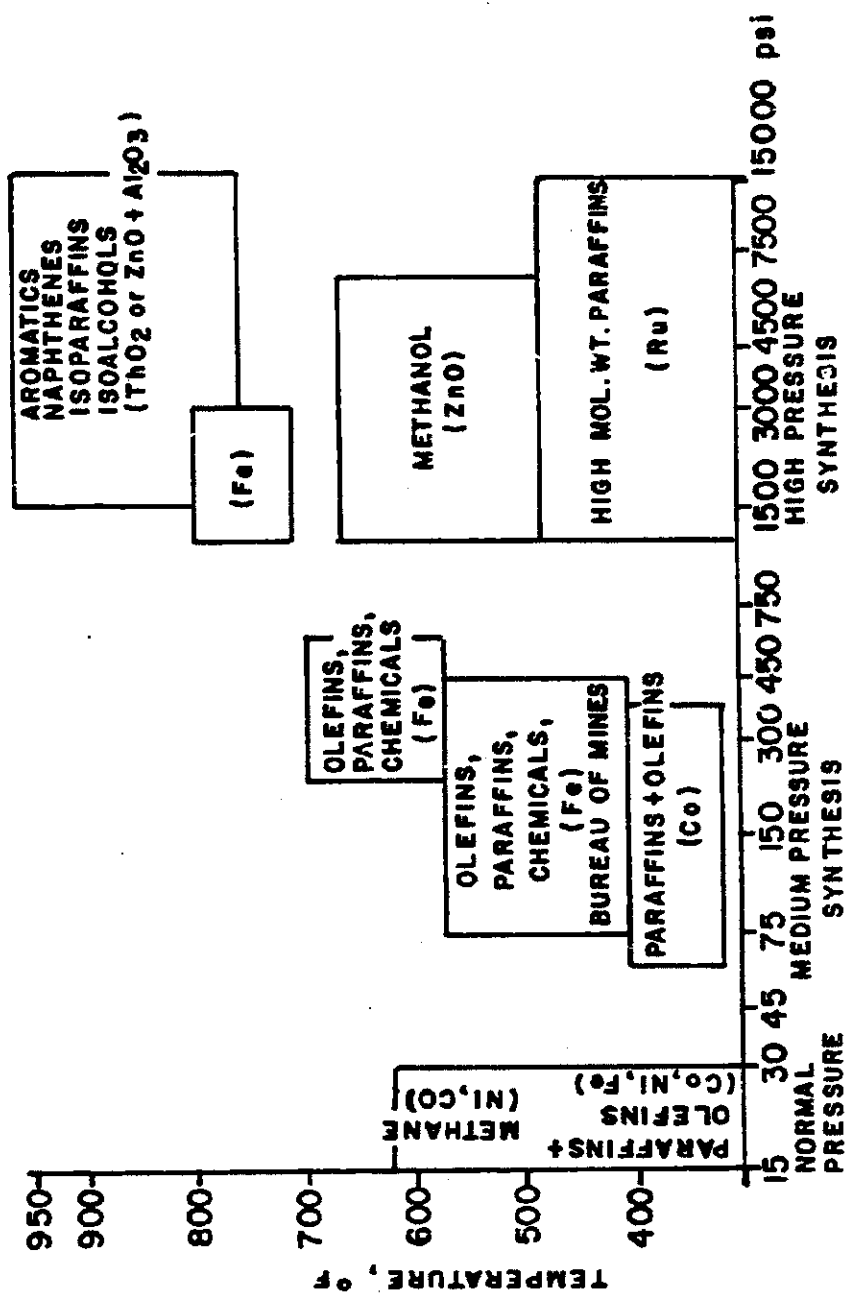


Fig. 3.3-1. Pressure-Temperature Regions for Synthesis

Table 3.3-2. Gasoline from Coal via Fischer-Tropsch Synthesis Using the Kellogg Synthesis Process (Synthol)⁹
Distribution of Products^a

Material	Product Rate, bbl/day
Gasoline	25,500
Diesel oil	1,230
Waxy oil	930
Propane-LPG	<u>2,000</u>
	29,660
Acetone	230
Methanol	30
Propanol	412
iso-Butanol	46
n-Butanol	136
MEK	56
n-Pentanol	<u>32</u>
	942
Aromatic solvent	158
Creosote	1,430
Coal tar fuel	1,860
Crude tar acids	<u>760</u>
	4,208
Sulfur	183 t/a
Ammonia	366 t/a

^aCoal feed rate 37,670 tons/day as received.
See Table 3.3-3, Case 1.

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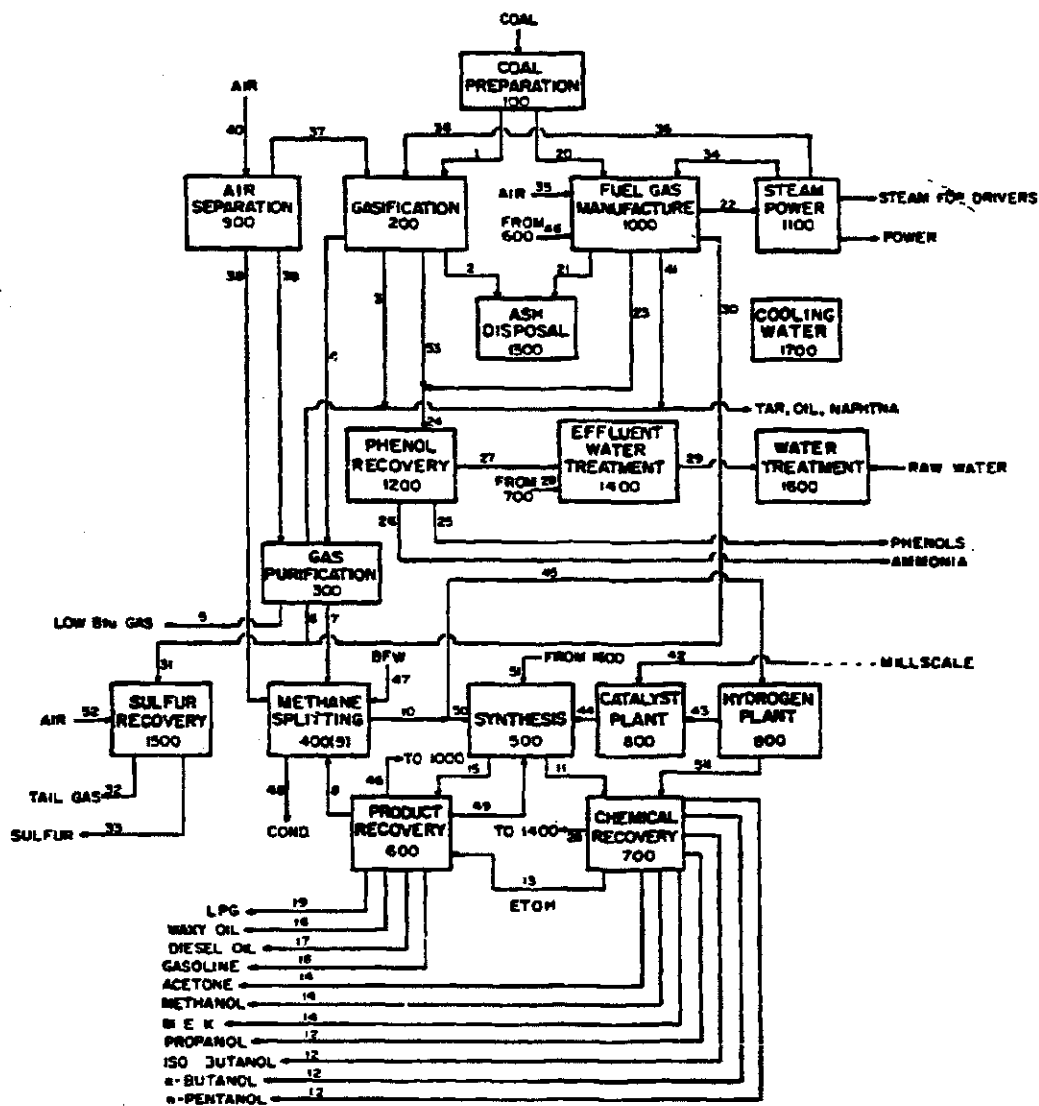


Fig. 3.3-2. Block Diagram for Manufacture of Gasoline by Fischer-Tropsch Synthesis Using the Kellogg Process.

Research and Development Needs. - Two principal steps in the process are coal gasification and Fischer-Tropsch synthesis which together directly and indirectly account for about 70% of the installed capital expenditure.

The cost of synthesis (including product recovery) is the smaller of the two, accounting for approximately 20% of the total cost. There appears to be very little in the way of improvements in this step of the process that may be anticipated from additional development effort. The yield of gasoline plus diesel oil (now about 77% of the total product) cannot be expected to improve significantly employing this complex synthesis reaction. Provision exists for recovery of heat produced in the synthesis reactor. A significant reduction in the cost of catalyst may not be expected. The iron catalyst, made from waste mill scale or iron ore, is the result of an extensive trial-and-error search.

In contrast, anticipated improvements in coal gasification will have a significantly favorable impact on the process. Kellogg's view is that Lurgi gasifiers must be used in any present-day design employing their synthesis method. These possess the following drawbacks:

1. Limited available size and throughput necessitate multiple parallel installation. Thus, SASOL has 9 Lurgi gasifiers in parallel, and the Kellogg's 25,500-bbl/day design has 30 parallel units each in the gasifier and fuel gas sections. Scale-up economies cannot be anticipated.
2. Only non-caking, bituminous coals may be used without pretreatment. This limits the use to western coals; eastern coals may be used with pretreatment; however, the exact nature of the pretreatment has not yet been determined.
3. Excess steam is employed in the feed to keep the ash below its clinkering temperature of about 1500°F. The excess steam is made from fuel gas manufactured on-site; hence the excess requirements represent addition of capital and operating costs involved in coal processing and fuel gas manufacture. Additionally, the excess steam is an added heat load in the gasifier which must be compensated for by oxidation of coal within the gasifier, thereby further increasing coal and oxygen usage. Finally, the excess steam contributes to phenol formation and hence requires a larger phenol recovery unit and larger effluent treatment facilities.

Kellogg estimates that a capital cost savings of about 35% could be achieved employing a second-generation gasifier such as the Synthane or Bi-Gas gasifiers instead of the Lurgi gasifier. Additionally, the process efficiency is expected to significantly improve due to an estimated reduction in coal feed of approximately 34%.

3.3.2 Characteristics of Fischer-Tropsch Gasoline Manufacturing Plants of Demonstration and Commercial Size

The following four cases have been considered:

- Case 1 - Demonstration plant producing 25,500 bbl/day of gasoline employing Lurgi gasifiers and purchasing deep-mined coal at \$7.50/ton.
- Case 2 - Commercial plant producing 100,000 bbl/day of gasoline employing Lurgi gasifiers and purchasing deep-mined coal at \$7.50/ton.
- Case 2A - Same as case 2 except that strip-mined coal at \$3.75/ton is assumed.
- Case 3 - Commercial plant employing a more efficient gasifier than the Lurgi and purchasing strip-mined coal at \$3.75/ton.

The results presume utilization of the Kellogg Synthesis Process shown in Fig. 3.3-2, and are based on a demonstration plant study performed by the M. W. Kellogg Company for EPA.⁹ Capital costs for the commercial size plant (Case 2) were estimated as follows: Those portions of the demonstration plant where economies of scale may be anticipated were scaled according to the 0.6 power law. Many sections of the plant (primarily the oxygen manufacturing section and the gasification section) consist of paralleled units of maximum available size. These were scaled proportionately. It is estimated that economies of scale are not available for approximately half the capital equipment items.

Discussions at the M. W. Kellogg Company clearly indicated the effect of inefficient gasification via the Lurgi gasifiers to be pervasive, affecting the cost of numerous capital items as well as the efficiency of the process, which is reflected by increases in the coal component of the operating costs. Thus

Case 3 is of interest even though it does not conform to present ground rules which presume use of presently-available gasifiers. Case 3 presumes use of either Bi-Gas or Synthane gasifiers which are being tested but are not yet commercially available. Kellogg estimates capital and operating cost savings of ~34% if second-generation gasifiers are used instead of Lurgi.

Thus Case 3 represents an estimate of the lowest realizable cost of gasoline via Fischer-Tropsch synthesis utilizing near-term but not presently commercial technology.

Raw materials and production rates are shown in Table 3.3-3. Estimated capital requirements for the four cases studied are given in Table 3.3-4. Estimated annual operating costs and unit production costs of gasoline are given in Table 3.3-5. All costs are presented in terms of 1973 dollars.

Table 3.3-3. Raw Materials, Utilities, and Products for Fischer-Tropsch Gasoline Demonstration and Commercial Plants

Case No.	1	2	3
Plant Type	Demonstration	Commercial	Commercial ^c
Coal Gasifier	Navaho Lurgi	Navaho Lurgi	Navaho New Gasifier
Raw Materials and Utilities			
As-received coal, tons/day	37,670	147,700	97,500
Dry coal, tons/day	31,460	123,400	81,400
Catalyst, ^a tons/day	77	302	302
Electricity			
Water consumption, ^b gpm	43,000	168,000	111,000
Energy input, 10 ⁹ Btu/day	662.2	2,597	1,714
Products			
Gasoline, bbl/day	25,500	100,000	100,000
Diesel oil	1,230	4,823	4,823
Waxy oil	930	3,647	3,647
LPG	2,000	7,843	7,843
Sub-total, bbl/day	29,660	116,313	116,313
Acetone, bbl/day	230	902	902
Methanol	30	118	118
Propanol	412	1,616	1,616
iso-butanol	46	180	180
n-butanol	136	533	533
MEK	56	220	220
n-pentanol	32	125	125
Sub-total, bbl/day	942	3,694	3,694
Aromatic solvent	158	620	620
Creosote	1,430	5,608	5,608
Coal tar fuel	1,860	7,294	7,294
Crude tar acids	760	2,890	2,890
Sub-total, bbl/day	4,208	16,412	16,412
Sulfur, tons/day	201	790	520
Ammonia, tons/day	366	1,435	990
Waste Streams			
Tail gas, tons/day	7,840	30,740	20,300
Ash, tons/day	7,670	30,100	19,900
Waste water, gpm	7,160	28,100	18,500
Process efficiency based on total coal ^c	22.1%	22.1%	33.5%
Process efficiency based on coal to gasifier ^d	50.4%	50.4%	~55%

^aMilli scale.^bEvaporation + process water.^cHeat of combustion of all hydrocarbons/EHV total coal.^dHeat of combustion of all hydrocarbons/EHV of coal to gasifiers.^eBased on an assumed reduction in coal consumption of 34%.

Table 3.3-4. Estimated Capital Requirements. Fischer-Tropsch
Gasoline Demonstration and Commercial Plants

Case No.	1	2	2A	3
Plant Type	Demonstration	Commercial	Commercial	Commercial ^b
Coal	Navaho	Navaho	Navaho	Navaho
Gasifier	Lurgi Gasifiers	Lurgi Gasif.	Lurgi Gasif.	New Gasifiers
Costs, \$ Millions				
On-Site				
Coal Preparation	34.55	78.42	78.42	Detail
Coal Gasification	39.24	153.82	153.82	not
Gas Purification	33.41	130.97	130.97	known
Methane Splitting	8.365	18.99	18.99	
Synthesis	43.41	170.17	170.17	
Product Recovery	23.00	52.20	52.20	
Chemical Recovery	6.535	14.83	14.83	
H ₂ & Catalyst Mfg.	6.951	15.78	15.78	
Oxygen Production	59.92	234.89	234.89	
Fuel Gas Mfg.	64.05	251.08	251.08	
Steam and Power	31.656	71.85	71.85	
Gas Liquor Treatment	10.92	42.81	42.81	
Ash Disposal	6.869	15.59	15.59	
Eff. Water Treatment	3.26	7.40	7.40	
Sulfur Recovery	7.40	16.79	16.79	
Raw Water Treatment	4.64	10.53	10.53	
Cooling Water	<u>21.875</u>	<u>49.65</u>	<u>49.65</u>	
On-Site Total	406.053	1435.76	1435.76	933.24
Off-Site and General	32.484	78.27	78.27	50.88
Engineering	43.854	151.40	151.40	98.41
Contingency	<u>65.751</u>	<u>227.10</u>	<u>227.10</u>	<u>147.61</u>
Total Plant Investment	548.142	1892.53	1892.53	1230.11
Interest during construction	87.70	302.80	302.80	196.82
Startup costs	27.33	101.62	57.76	41.64
Working capital	<u>17.93</u>	<u>70.3</u>	<u>39.85</u>	<u>29.00</u>
Total Capital Requirement	681.10	2367.25	2292.94	1497.57

^aAll costs are based on 1973 dollars with no allowance for escalation.

^bRoughly estimated.

Table 3.3-5. Estimated Annual Operating Costs and Fixed Charges for Demonstration and Commercial-Size Fischer-Tropsch Gasoline Production

Case No.	1	2	2A	3
Gasoline Production, bbl/day	25,500	100,000	100,000	100,000
Type Plant	Demonstration	Commercial	Commercial	Commercial
Type Coal	Deep-Mined	Deep-Mined	Strip-Mined	Strip-Mined
Type Gasifier	Lurgi	Lurgi	Lurgi	New Gasifier
Annual Costs, \$ Millions				
Raw Materials & Utilities				
As received coal ^a	93.2	365.5	146.22	96.53
Electricity ^b	-	-	-	-
Catalyst	0.25	1.00	1.00	1.00
Treated water	2.04	7.98	7.98	5.27
Labor and Administration				
Operation	8.3	24.77	24.77	24.77
Maintenance	8.22	28.39	28.39	18.45
Supervision	2.49	7.97	7.97	6.48
Administration	11.42	36.69	36.69	29.82
Supplies				
Operating	2.49	7.43	7.43	7.43
Maintenance	8.22	28.39	28.39	18.45
By-Product Credit				
Ammonia @ \$25/ton	(3.02)	(11.83)	(11.83)	(7.81)
Diesel oil @ 15¢/gal	(2.56)	(10.04)	(10.04)	(10.04)
Heavy oil @ 30¢/MMBtu	(0.52)	(2.04)	(2.04)	(2.04)
LPG @ \$1.20/MMBtu	<u>(3.24)</u>	<u>(12.69)</u>	<u>(12.69)</u>	<u>(12.69)</u>
Ann. Operating Cost	127.29	471.52	252.19	175.62
Fixed Charges @ 23.4%	159.38	553.94	536.55	350.43
Total Capital Requirement	<u> </u>	<u> </u>	<u> </u>	<u> </u>
Total Operating + Fixed Charges	286.67	1025.46	788.74	526.05
Production Cost of Gasoline				
\$/10 ⁶ Btu	6.46	5.89	4.53	3.02
\$/bbl	34.06	31.07	23.90	15.94
¢/gal	81.1	74.0	56.9	38.0

^aDeep-mined coal @ \$7.50/ton; strip-mined coal @ \$3.00/ton.

^bAll power produced on-site. Air compressors driven by steam.

3.3.3 References

1. Report of the Committee on Coal Derivatives, British Ministry of Power Report, August 1960.
2. W. B. Johnson, "Coal Beats Oil Here," Petroleum Refiner 35, 222-228 (1956).
3. "Shut Down Major Chemical, Synthetic Gasoline Plant," Ch.E. Prog. 53, 50 (1957).
4. H. C. Hottel, J. B. Howard, New Energy Technology, The MIT Press (1971).
5. Kirk-Othmer, Encyclopedia of Chemical Technology, Vol. 4, Second Ed. (1964).
6. H. H. Lowrie, Chemistry of Coal Utilization, Table 24, p. 990, John Wiley & Sons, New York, 1963.
7. W. A. Gruse, D. R. Stevens, Chemical Technology of Petroleum, McGraw-Hill, Third Ed. (1960), pp. 636-642.
8. H. Tramm, "25-yrs of F/T Synthesis," 5th World Petrol. Cong., Sect. III, New York (1959), pp. 347-361.
9. M. W. Kellogg Co., Gasoline from Coal Via the Fischer-Tropsch Synthesis, Study 68-02-1308 for EPA, January 1974.
10. H. H. Storch et al., The Fischer-Tropsch and Related Syntheses, John Wiley & Sons, New York (1951).
11. The Bureau of Mines Synthetic Liquid Fuels Program, Report of Investigations 5506, Dept. of Interior (1959).

3.4 Methanol from Coal

The manufacture of methanol¹ from coal involves the three principal operations of (1) coal gasification to produce a raw synthesis gas, (2) adjustment of the composition of the synthesis gas to yield the proper stoichiometric ratio of hydrogen to carbon monoxide (2:1), and (3) production of methanol from CO and H₂ by catalytic synthesis (Fig. 3.4-1). The sulfur (as H₂S) released during the gasification of coal and the excess CO₂ produced during CO shift conversion are removed immediately before methanol synthesis. In the following, a brief description is given of the existing commercial technology for each of the operations shown in Fig. 3.4-1. It should be noted that intensive efforts are under way to develop new processes for coal gasification and, to a lesser extent, methanol synthesis. However, in keeping with the objectives of this project to demonstrate the use of existing technology, subsequent discussion is limited to commercially proven processes.

3.4.1 Coal Gasification Processes

Three commercially proven processes for the gasification of coal, which basically differ only in the design and operating conditions of the gasifier, are the Koppers-Totzek, Winkler, and Lurgi processes. The traditional use of the product gas from these processes has been either as an intermediate-Btu fuel for industrial heating or power production or as a synthesis gas for the manufacture of ammonia. Coal-produced gas has not been used for methanol production in the U.S. because of the availability of low-cost natural gas.

Koppers-Totzek. - The Koppers-Totzek gasifier is a refractory lined steel shell wherein oxygen, steam, and pulverized coal (70% at 200 mesh) are brought together in opposing burner heads spaced 180° (2 heads) or 90° (4 heads) apart.³ Although no gasifiers of the latter design have as yet been commercially tested, two commercial plants now under construction by Koppers will use them exclusively. Upon entering the gasifier at approximately atmospheric pressure the reaction mixture is gasified immediately at temperatures on the order of 1650°C. Gaseous hydrocarbons released from

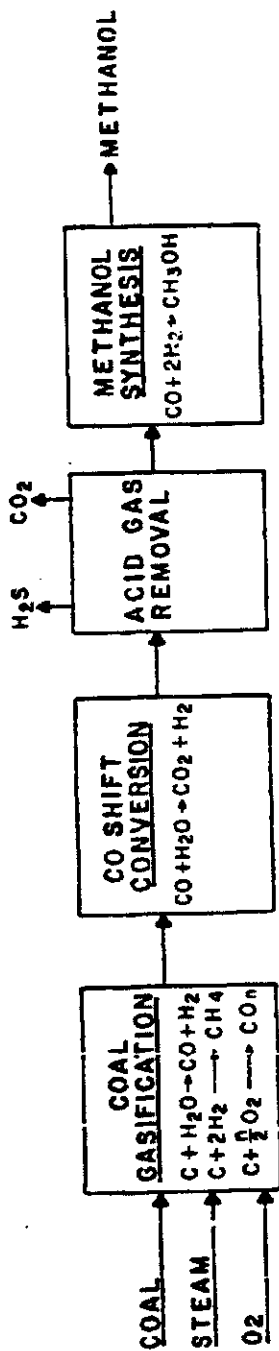


Fig. 3.1-1. A Generalized Flowsheet for the Manufacture of Methanol from Coal.

the coal are rapidly decomposed to CO, CO₂, and H₂. At this high temperature no coal liquids are formed, and approximately 50% of the ash is fused to a molten slag which drops into a quench tank under the gasifier. The remainder of the ash is recovered downstream.

With respect to the suitability of coal feedstocks for gasification, the Koppers-Totzek gasifier is the most versatile of the three gasifiers under consideration. Its high temperature operation allows it to process coal of any rank in addition to a variety of hydrocarbon liquids and gases. However, such operation may require paying a severe penalty in oxygen consumption, although this can be somewhat compensated for by an increased production of process steam in the shell of the gasifier. The low pressure operation of the gasifier also provides for relatively uncomplicated operating equipment and maintenance, but this does necessitate more downstream compression of synthesis gas.

Winkler. - Gasification of coal (-3/8 in.) in the Winkler gasifier takes place in a bed fluidized by injected oxygen and steam.⁴ Although the operating temperature of the fluid bed ranges from 800-1000°C, depending on the ash fusion temperature of the coal, even higher temperatures are reached above the bed during a second gasification initiated by an additional injection of oxygen and steam. This second reaction zone is needed to gasify the unreacted carbon entrained in the large amount of ash (~70%) which escapes from the fluid bed. It also prevents formation of any liquid products and limits CH₄ production to 3% (volumetric, dry). A radiant boiler cools the gas approximately 400°F before leaving the generator, thereby preventing ash particles from melting on the refractory walls of the exit ducts. The operating pressure of Winkler gasifiers in existing commercial installations is 1.5 atm, but operation at 3 atm has been stated to be feasible.*

Although some bituminous coals have been processed, the Winkler gasifier was designed primarily to handle noncaking, high-volatile lignite. The relatively high reactivity of this type of coal helps to minimize the escape of unreacted carbon from the secondary gasification zone. Satisfactory

* Private communication, Davy Powergas, Inc., Lakeland, Florida, Jan. 8, 1971

carbon utilization in this zone also requires careful flow control on the secondary injection of oxygen and steam. In general, the operation of the Winkler gasifier appears somewhat more complicated than that of the Koppers-Totzek gasifier, but it does result in reduced oxygen consumption. It can be anticipated that the operation and maintenance of a pressurized Winkler would be even more complex, but this might, of course, be compensated for by improved performance.

Lurgi. - The Lurgi gasifier is a reactor for the countercurrent gasification of coal (1/4 - 1-1/2 in.) in a moving bed at pressures ranging from 20 to 30 atm.⁵ This pressurized operation makes Lurgi the most complicated of the three gasifiers to operate and maintain. External, pressurized lock hoppers for both coal feed and ash withdrawal are required in addition to numerous moving internal parts. However, the elevated pressure does permit gasification temperatures to be lowered to a maximum of approximately 800°C with a residence time of about one hour. In addition to increasing methane production to approximately 10% (volumetric, dry), this reduced temperature results in formation of liquids such as tars, heavy oils, and phenols which must be removed in a scrubbing cooler. Opinions on the desirability of producing these materials vary. Some regard them as valuable by-products to be marketed, or as medium Btu fuel to compensate for the reduced production of process steam due to the relatively low operating temperature of the gasifier. Others consider the added processing and handling of these materials as very undesirable complications, particularly if the overall objective is to produce chemical synthesis gas.

Of the three gasifiers under consideration, the Lurgi is also the most restrictive with regard to acceptable characteristics of the coal feedstock. In addition to the constraint on particle size indicated above, the swelling index and moisture content of Lurgi coals generally must be less than 4.5 and 20%, respectively, in order to ensure satisfactory operation of the moving bed of coal. Thus, Lurgi gasifiers have traditionally processed only sized, noncaking coals. However, experiments are now under way to test, and if necessary to modify, the Lurgi gasifier to process caking American coals.⁶ Successful results have already been achieved with the mildly caking Illinois No. 6.

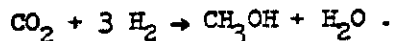
3.4.2 Commercial Methanol Synthesis Processes

One common method for classifying processes for conversion of CO and H₂ to methanol is as a high pressure or a low pressure process.⁷ Basic flowsheets for the two processes are essentially the same; the difference in operating pressure is related to the type of catalyst used. The older high pressure process uses a catalyst containing zinc oxide and chromic oxide in various proportions with or without other metal oxides as promoters. The activity of the zinc oxide-chromic oxide catalyst with or without other metal oxides shows a maximum at temperatures of about 350°C. The activity is such that in order to yield adequate single-pass conversions under prevailing reactor conditions, the reactor pressure must be in the range of 200-350 atm. In 1967, Imperial Chemical Industries (ICI) introduced a copper-based catalyst specially designed for methanol synthesis at or below 100 atm and 300°C. The activity of the conventional zinc-chromium oxide catalyst would be unacceptable at these conditions. In addition, this copper-based catalyst is more selective for methanol than zinc-chromium formulations, but the latter are more tolerant of sulfur (~3-5 ppm) in the synthesis gas.

The manufacture of methanol starts with the preparation of a synthesis gas such as that produced by the gasification of coal. The feed to the converter, designated as the makeup gas, is a combination of synthesis gas and recycle gas from a high pressure separator. The composition of this makeup gas varies with the design, but its overall composition must be adjusted to make the ratio H₂/(CO + 1.5 CO₂) slightly greater than 2.0. The converter effluent is somewhat cooled by heat exchange with the makeup gas, is further cooled to condense the methanol product, and then enters a separator vessel. A portion of the gas from the separator is purged to remove impurities and inerts introduced with the synthesis gas or produced from side reactions in the converter. The remainder of the gas from the separator is recycled to the converter. The condensed crude methanol is dewatered in an atmospheric distillation tower. Depending on the desired purity in the final product, additional refining of the crude methanol to remove higher alcohols may be required. For use as an automotive fuel, the higher alcohols are desirable and need not be removed.

Temperature control of the exothermic methanol synthesis reaction is essential. As shown in the schematic flow diagram, one method of temperature control is to introduce portions of cold makeup gas

directly into the converter at several points, thus providing a quench or "cold shot" type of cooling. Additional temperature control is provided if CO_2 is present in the synthesis gas. CO_2 is converted to methanol according to the following equation:



However, the exothermic heat of this reaction is less than that released by the conversion of CO. Thus, the average rate of heat generation if both carbon oxides are present in the feed is less than that when pure CO is used. Temperature moderation is particularly important for use with the copper catalyst since it is much more susceptible to sintering than the zinc-chromium oxide catalyst.

3.4.3 CO Shift Conversion and Gas Purification

The composition of the gas leaving the gasification section is determined by the operating conditions of the gasifier. The technology for adjusting the H_2/CO ratio of this gas for subsequent processing is well developed. It is based on the CO shift conversion reaction, using additional steam to convert CO to H_2 and CO_2 . Early catalysts for shift conversion consisted of iron oxides with chromium promoters for operation at high temperatures.⁸ Low temperature catalysts were later developed by replacing the chromic oxides with those of copper and zinc. In either case the activity of the iron oxides was substantially reduced by H_2S in the synthesis gas. The invention of catalysts containing sulfided cobalt and molybdenum oxides, however, has eliminated the need for upstream desulfurization of the synthesis gas.⁹ This switch provides for better conservation of sensible heat during shift conversion, and reduces the requirement for make-up steam.

There are several acid-gas removal systems on the market today that could be used for the purification of coal-derived gases. The estimates in this section are based on the use of the Rectisol process.¹⁰ This process utilizes a methanol solvent at approximately -40°C to remove both H_2S and CO_2 from the gas stream. Regeneration of the solvent yields a CO_2 stream which may be vented and an H_2S stream which goes to a conventional Claus plant for sulfur recovery.¹¹

3.4.4 Conceptual Designs of Methanol-from-Coal Plants

Twenty-two of the twenty-five methanol plants planned or built worldwide since 1967 are under license from Imperial Chemical Industries, all of them with the Cu catalyst for operation at low temperature and pressure. The last high pressure methanol synthesis plant was built in 1968 by DuPont at Beaumont, Texas. In this study of possible processes for manufacturing methanol from coal, the low pressure process for methanol synthesis was selected because it appeared to be the one preferred by most recent methanol manufacturers. In addition, the low pressure process has certain technical advantages over the high pressure process due to its lower compression requirements and capital investment. However, it should be pointed out that a greater percentage of higher alcohols (C_2-C_4) can be produced with the high pressure process because the zinc-chromium catalyst, as previously discussed, is less selective for methanol. The higher alcohols are desirable components in blends of methanol and gasoline because they increase the amount of water that can be tolerated without phase separation.

Conceptual flowsheets for each of the three gasifiers, combined with a low pressure methanol synthesis plant, have been developed. One type of alternative represents the combination of either a Koppers-Totzek gasifier or a Winkler gasifier with a methanol synthesis step. The sequence for these cases is gasification, followed by gas cleanup. The particulate-free gas is fed to a shift converter where additional hydrogen is formed through the reaction of CO and steam. The gas from the shift converter is sent to a purification step (e.g., Rectisol) where all the H_2S and most of the CO_2 is removed. The gas coming from the acid gas removal step is sent to methanol synthesis.

Because the Lurgi gasifier produces approximately 10% methane, it is proposed that a Lurgi based methanol synthesis process manufacture methanol and substitute natural gas (SNG) as co-products. This type of flowsheet provides for crude gas cleanup and acid gas removal before a single pass through a methanol converter. Elimination of the conventional methanol recycle loop also permits delaying CO shift until after methanol is removed

as a product. Following a second acid gas removal treatment, a 30% CH₄ feedstock is upgraded to SNG by catalytic methanation. This last processing step may not be considered commercially proven at 100 atm, but current development efforts are so intense that it is anticipated to be so in the near future.

In order to compare these three proposed flowsheets, material balances were derived for production of 5000 tons/day of methanol. It is a general consensus that, for a single module of production, this capacity will fully utilize all the required equipment, including in particular a single train of compressors. These material balances are shown in Table 3.4-1 along with an estimate of the additional coal required to produce enough intermediate Btu fuel gas to provide all the power required for utilities and off-sites. Cases 1, 2, and 3 represent conceptual demonstration plants using the Koppers-Totzek and Winkler gasifiers. Case 4 represents a conceptual commercial plant using a Koppers-Totzek gasifier and a western coal. Case 5 represents a conceptual commercial plant using a Lurgi gasifier and a western coal for co-production of 5000 tons/day of methanol and 185 million ft³/day of SNG.

Estimates of the required capital investment for these plants are presented in Table 3.4-2. The estimated operating costs, revenues, and unit production are shown in Table 3.4-3. Manpower and steel requirements for the Case 5 plant for co-production of methanol and SNG are shown in Table 3.4-4.

In all three of these processes, sulfur is a by-product. No credit has been taken for this production, however, because of the uncertainty in its future supply and demand. Credits for other materials also were not taken. For example, the plant in Case 5 produces coal tars, phenols, etc., which are probably more valuable as chemical raw materials than as the fuel for which they were considered. Large quantities of carbon dioxide are produced in all three cases. There is a potential demand for CO₂ in large quantities as a pressurizing medium for secondary recovery in oil fields if it is supplied under pressure. However, such a market for this material is also uncertain at this time.

Table 3.4-1. Estimated Raw Materials, Power, Products, and Effluents for the Manufacture of Methanol from Coal using the ICI Methanol Synthesis Process and the Indicated Gasifier^a

Case No.	1	2	3	4	5
Plant Type	Demonstration	Demonstration	Demonstration	Commercial	Commercial
Coal Type	Clifty Creek #6	Lincoln Co. Wyo.	Navaho	Lincoln Co. Wyo.	Navaho
Gasifier Type	Koppers-Totzek	Koppers-Totzek	Winkler	Koppers-Totzek	Lurgi
Raw Materials					
As-received coal, t/d	8,260	8,500	11,690	34,000	20,670
Dry coal, t/d	7,210	6,570	9,760	26,300	17,260
Oxygen, t/d (produced on-site)	6,700	6,860	5,000	27,440	5,500
Water, gpm	3,300	3,700	2,900	14,800	6,200
Electricity, kW	^a 177	^a 181	^a 202	^a 724	^a 367
Energy input, 10 ⁹ Btu	177	181	202	724	367
Power for Utilities and Off-Sites					
As-received coal, t/d ^b	1,650	1,700	2,040	6,800	5,280
Energy input, 10 ⁹ Btu/d	35	36	35	144	94
Products and Effluents					
Methanol (@ 9,770 Btu/lb), t/d	5,000	5,000	5,000	20,000	5,000
Substitute natural gas (@ 950 Btu/scf), 10 ⁶ scf/d	0	0	0	0	185
Total energy output, 10 ⁹ Btu/d	97.7	97.7	97.7	391	273 ^c
Sulfur, t/d	390	90	100	360	148
Carbon dioxide, t/d	9,150	-	-	-	-
Ammonia, t/d	0	-	-	-	200
Waste water, gpm	210	3,000	2,800	12,000	-
Solid wastes, t/d	1,260	510	6,100	2,040	4,730
Tars, oils, and phenols	0	0	0	0	1,860
Process Efficiency					
Total energy output/total energy input, %	46	45	41	45	59 ^c

^a Electricity is generated on-site from a portion of coal feed as specified under "Power for Utilities and Off-Sites."

^b As-received coal for power generation is gasified and desulfurized before combustion.

Table 3.4-2. Estimated Capital Investment for Conceptual Methanol from Coal Plants
(Millions of 1973 Dollars)

Case No.	1	2	3	4	5
Plant Type	Demonstration	Demonstration	Demonstration	Commercial	Commercial
Coal Type	Clifty Creek #6	Lincoln Co. Wyo.	Navaho	Lincoln Co. Wyo.	Navaho
Gasifier	Koppers-Totzek	Koppers-Totzek	Winkler	Koppers-Totzek	Lurgi
On-Site Process Units	157.3	156.4	150.1	453.6	235.5
Off-Sites and Utilities -	28.6	28.6	28.6	82.5	151.1
Initial Catalysts and Chemicals	a	a	a	a	a
Contingency	27.9	27.8	26.8	80.4	58.0
Start-up Costs	8.4	5.5	5.7	17.9	12.0
Interest During Construction	34.5	34.3	33.2	99.4	71.8
Working Capital	4.7	4.7	4.7	13.5	9.8
Total Capital Requirement	261.4	252.6	249.1	747.3	528.4

^aIncluded in cost of on-site process units.

Table 3.4-3. Estimated Annual Costs for the Manufacture of Methanol from Coal Using the ICI Methanol Synthesis Process and the Indicated Gasifier

Case No.	1	2	3	4	5
Plant Type	Demonstration	Demonstration	Demonstration	Commercial	Commercial
Coal Type	Clifty Creek #6	Lincoln Co. Wyo.	Navaho	Lincoln Co. Wyo.	Navaho
Gasifier Type	Koppers-Totzek	Koppers-Totzek	Winkler	Koppers-Totzek	Lurgi
Annual Costs, \$10 ⁶					
Direct Labor					
Labor	2.89	2.89	2.17	4.68	3.50
Supervision	0.96	0.96	0.71	1.55	1.65
Maintenance	3.50	3.50	3.56	9.87	7.47
Payroll Overhead	4.40	4.40	3.26	9.64	7.57
Supplies	4.37	4.37	3.91	17.5	8.52
Cost of As-Received Coal	24.53	10.10	13.59	40.39	25.69
Catalysts and Chemicals	0.88	0.88	0.88	3.52	4.52
Water	0.47	0.53	0.41	2.12	0.88
Capital Charges @ 23.4%	61.2	59.1	58.3	174.9	123.6
By-Product Credits	0	0	0	0	(13.31)
Total Annual Cost, \$10 ⁶	103.2	86.7	86.8	264.2	170.1
Total Energy Production, 10 ¹² Btu/yr	32.24	32.24	32.24	129.0	90.09
Methanol Production Cost					
\$/10 ⁶ Btu	3.20	2.69	2.69	2.05	1.89 ^a
\$/bb1	8.70	7.31	7.32	5.57	5.13
\$/gal ^b	20.7	17.4	17.4	13.3	12.2

^aMethanol and SNG assumed to be produced at the same unit cost per million Btu.

^b64,700 Btu/gal.

Table 3.4-4. Estimated Manpower and Steel Requirements for a Case 3 Plant Producing 185 MM SCFD SNG and 5000 tons/day Methanol from Coal^a

Item	Requirements
Gasification Plant	
Home Office, man-hour	1,320,000
Field Construction, man-hour	6,830,000
Shop Fabrication, man-hour	5,000,000
Operations, man-year	612
Steel, tons	50,000 - 60,000
Coal Mine	
Home Office, man-hour	200,000
Field Construction, man-hour	700,000
Shop Fabrication, man-hour	2,400,000
Steel, tons	40,000
Methanol Synthesis Plant	
Home Office, man-hour	250,000
Field Construction, man-hour	1,356,000
Shop Fabrication	750,000
Operations, man-year	80
Steel, tons	7,500

3.4.5 Other Considerations

For a first demonstration plant, thermal efficiency and economics are not overriding considerations for choosing a process. More important considerations are felt to be the length of time required to design, construct, and commission the plant, and the assurance that reliable operation can be obtained. To conserve coal, it is desirable that the thermal efficiency of a commercial process for converting coal to methanol be as high as possible. However, maximum thermal efficiency does not necessarily mean minimum product cost.

The capacity recommended for the demonstration plant, 5000 tons of methanol per day, requires some comment. At present, there is no 5000-ton/day methanol plant in operation, either single-train or multiple train. However, all components of such a plant have been reported to be operating under conditions similar to those experienced in methanol production. There are no integrated methanol-from-coal plants in operation, but the technology used in existing ammonia-from-coal plants is similar and has been demonstrated.

The coal used in a large methanol-from-coal plant will probably be obtained from a mine close to the plant site. Although all of the gasifiers considered are commercially available, the selection of the gasifier depends on the type of coal available at a particular site. For this reason, all of the gasifiers discussed are potentially eligible for use in the demonstration plant.

3.4.6 References

1. T. B. Reed and R. M. Lerne "Methanol: A Versatile Fuel for Immediate Use," Science 182 (4119), 1299-1304 (1973).
2. A. L. Conn, "Low Btu Gas for Power Plants," Chem. Eng. Progr. 69 (12), 56-61 (1973).
3. J. F. Farnsworth, H. F. Leonard, D. M. Mitsak, and R. Wintrell, The Production of Gas from Coal Through a Commercially Proven Process, Koppers Company, Inc., Pittsburgh, Pa., August 1973.
4. I. N. Banchik, The Winkler Process for the Production of Low Btu Gas from Coal, Davy Powergas, Inc., Lakeland, Florida (1973).
5. P. F. H. Rudolph, "The Lurgi Process: The Route to SNG from Coal," Proceedings of the Fourth Synthetic Pipeline Gas Symposium, AGA Catalogue No. L 11173, 175-214 (October 1972).
6. D. C. Elgin and H. R. Perks, "Trials of American Coals in the Lurgi Pressure-Gasification Plant at Westfield, Scotland," Proceedings of the Fifth Synthetic Pipeline Gas Symposium, American Gas Association, in press.
7. S. Strelzoff, "Methanol: Its Technology and Economics," in G. A. Danner (ed.), "Methanol Technology and Economics," Chem. Eng. Progr. Sym. Ser. No. 98, 66, 54-68 (1970).
8. A. P. Ting and S. W. Wan, "Sizing CO Shift Converters," Chemical Engineering, 185-192 (May 19, 1969).
9. E. C. Schora, "Other Coal Gasification Related A.G.A. Programs at the Institute of Gas Technology," Proceedings of the Fifth Synthetic Pipeline Gas Symposium, American Gas Association, in press.
10. G. Hochgesand, "Rectisol and Purisol," Ind. Eng. Chem. 62 (7), 37 (1970).
11. C. S. Barry, "Reduce Claus Sulfur Emissions," Hydrocarbon Processing 51 (4), 102 (1972).

4. PRODUCTION OF SUBSTITUTE NATURAL GAS

The Western Gasification Company (WESCO) and the El Paso Natural Gas Company (EPNG) currently are seeking authorizations to complete engineering and construct 250 million standard ft³/day SNG plants in Northwest New Mexico. Initial production of SNG in these plants is expected in late 1977. Other similar commercial plants are planned by these companies as well as several other gas utility companies.

The WESCO and EPNG plants will utilize the Lurgi high-pressure coal gasification process. Since 1936, the Lurgi process has been used commercially in 19 plants (50 gasifier units) in many parts of the world, including Germany, Scotland, South Africa, and South Korea. Adaptation of the basic Lurgi technology requires the addition of a "methanation" step (catalytic reaction to convert CO to CH₄) to produce 920-950 Btu/scf CO-free gas required in the United States. Although there are no current commercial plants practicing large-scale methanation of coal gas, the technology of methanation has been widely practiced in other applications. Lurgi engineers have made extensive studies of coal gas methanation technology and catalysis on a pilot-plant scale. They are confident that the successful extension of these studies to commercial practice is technically and economically feasible at this time. To supplement Lurgi work in this area, WESCO, in combination with several major gas companies, has undertaken independent pilot-plant and engineering studies for large-scale coal gas methanation.

The following presents a capsule description of certain features of these projects, including: estimated raw materials, utilities, and products (Table 4-1); a flow diagram and material balance around the gasification system (Fig. 4-1); a flow diagram and material balance for water (Fig. 4-2); a diagram and characteristics of the sulfur disposition systems (Fig. 4-3); estimates of the required capital investment (Table 4-2); and estimates of annual expenses and revenues (Table 4-3).

The advice of the Fluor Corporation is that a total of 8 years should be allowed for the steps of (1) conceptual design, (2) Title I engineering

Table 4-1. Estimated Raw Materials, Utilities, and Products of Commercial Plants for Producing SNG from Coal^a

	WESCO	El Paso Natural Gas
Raw Materials and Utilities		
As-received coal - gasifiers, tons/day	21,860	25,947
- steam boiler, tons/day	3,760	-
Electricity, kW	28,500	-
Water, gpm	5,100	6,200
Total energy, 10 ⁹ Btu	440	460
Output		
High-Btu gas, 10 ⁶ ft ³ /day	250	250
, 10 ⁹ Btu/day	238	238
, tons/day	5,440	5,500
Sulfur, tons/day	174	148
Phenols, tons/day	105	122
Tars, oils, and naphtha, tons/day	1,475	1,737
Off-gas, tons/day	792	
CO ₂ gas, tons/day	16,631	
Ash (dry basis), tons/day	6,433	4,730
Water, gpm		
to atmosphere	3,550	4,700
to mine with ash	1,030	150
Process Efficiency, % ^b	67	63

^aWestern strip-mined coal.

^bEfficiency is expressed as thermal energy of SNG product divided by thermal energy of coal feed.

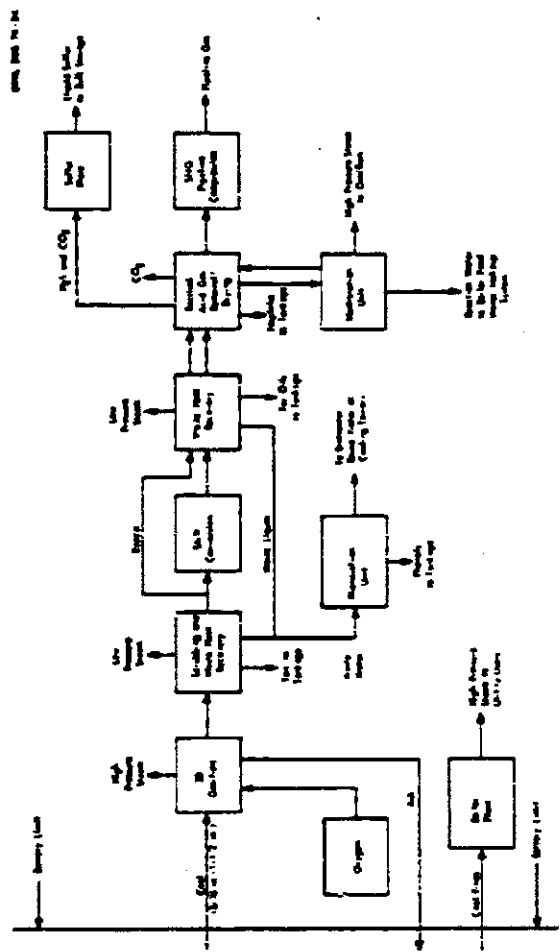
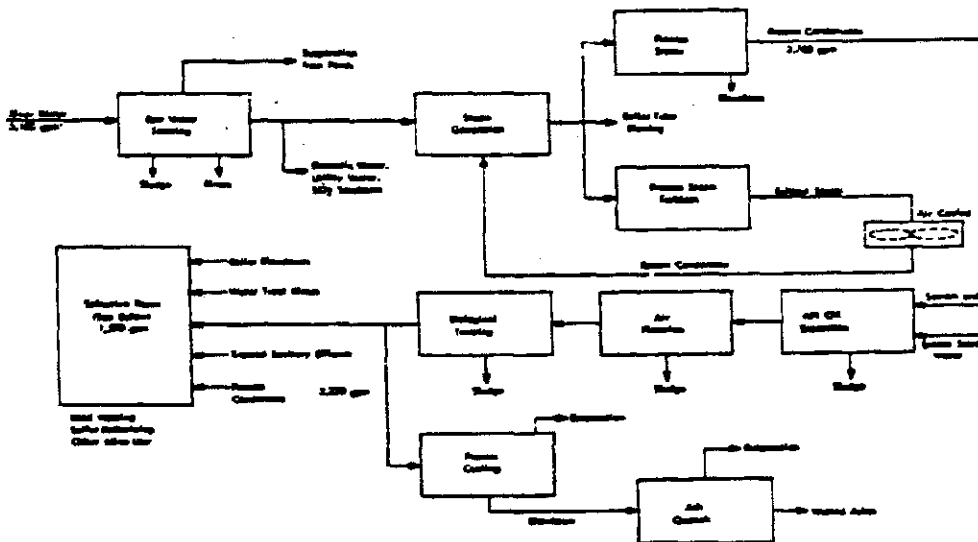


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Category 1	Category 2	Category 3	Category 4
100	200	300	400
500	600	700	800
900	1000	1100	1200
1300	1400	1500	1600
1700	1800	1900	2000
2100	2200	2300	2400
2500	2600	2700	2800
2900	3000	3100	3200
3300	3400	3500	3600
3700	3800	3900	4000

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WTR, 200 70-20



* 0.250 gpm/ft² area

Water Requirements and Discharges

	gpm	ft ³ /day
Process Consumption		
To supply buildings	1,750	
Returned to atmosphere evaporated	1,750	
Net consumption	0	0.2
Water to Atmosphere		
Evaporation		
From raw water tank	40	
From cooling tower	1,750	
From evaporating tank unit	150	
From cooling-tower make-up	170	
From cooling of final water	1,730	
	3,840	
Net bleed water		
From 2000 capacity of hot-water tank	200	
From 2000 gal. 10% solution	20	
	220	
Total return to atmosphere	3,720	0.7

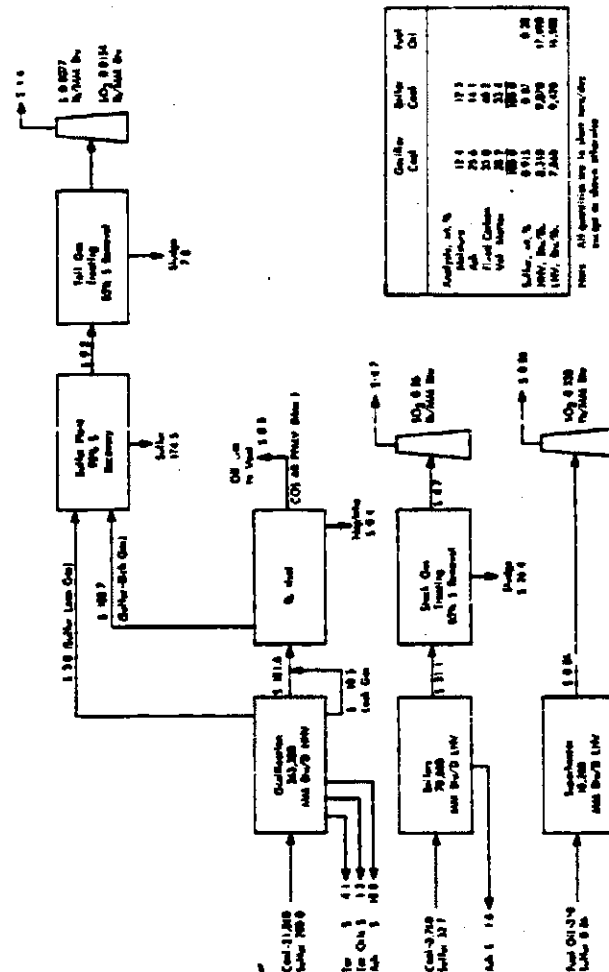
	gpm	ft ³ /day
Sludge to other		
To water treating sludge	50	
To water treating tank	20	
To water treating tank	20	
Total sludge to other	90	0.2
Other		
Returned to hot water	20	
Returned to hot water	20	
Total other	40	0.1
Grand Total	3,860	0.9

* Does not include water demand from heating of boiler fuel

Source: "Cool Condensate - A Technical Handbook" by James G. Gorman, Chemical Engineers, New York

Fig. 4-2. Water Treatment and Reuse Systems.

Sheet No. 11-16



Analysis, %	Gasifier Gas	Shift Gas	Gas Out
Hydrogen	12.4	14.1	17.3
Carbon Monoxide	28.6	24.1	22.1
Carbon Dioxide	32.9	32.5	32.5
Water Vapor	19.8	19.8	19.8
Sulfur, %	0.015	0.01	0.01
HW, %	0.210	0.210	0.210
HW, %	7.800	7.800	7.800

Note: All quantities are in scfm based on gas at 2000 ft. pressure.

With the two stage processing of the H₂S-bearing gas stream during the normal operation, 1.5 tons/day of sulfur will be produced and 1.5 tons/day of sulfur dioxide will be produced. This amount is 0.007% of sulfur dioxide of the feed gas stream.

The Claus plant design is based on two trains, three operating and one standby, each with three reaction stages, followed by a common tail gas treating unit with standby equipment.

The tail gas scrubber will have a flow volume of 70,000 scfm/day, when the normal 3,200 tons/day of tail gas is scrubbed. The scrubbing system will remove 85% of the sulfur from the tail gas stream. This, also, means that 2,680 tons/day of sulfur dioxide will be added to the tail gas stream. This, also, means that 2,680 tons/day of sulfur dioxide will be added to the tail gas stream.

As indicated above, the total sulfur contained in the tail gas to the large gasifier will be 200 tons/day. Of this, 180 tons, will be captured by the gasifier and 20 tons will be captured by the shift converter.

The remaining sulfur distribution is as follows by product gas stream is as follows:

- CO₂ - 100 tons/day
- CO₂ - 100 tons/day
- CO₂ - 100 tons/day
- Total - 300 tons/day

Of this total, 100 tons/day of operations will be automatically generated in the Claus sulfur plant, 100 tons/day will be generated in the shift converter. The Claus plant will remove the sulfur from the tail gas stream. The remaining tail gas from the plant will be further processed in a gas scrubbing system. This will remove 85% of the sulfur from the tail gas stream.

Source: Coal Gasification - A Technical Description
 American Gasification Company
 Lexington, Massachusetts

Table 4-2. Estimated Capital Investment (1973 Dollars) for the WESCO and El Paso Natural Gas Plants for Production of Synthetic Natural Gas from Coal

	Capital Cost, \$ millions	
	WESCO	EPNG
Process Units		
Gas Production		64.45
Crude Gas Shift Conversion		7.68
Gas Cooling and Purification		48.72
Methane Synthesis		18.55
Product Compression and Dehydration		5.66
Liquids Separation and Purification		18.65
Lock Gas Storage and Compression		1.92
Sulfur Recovery		8.16
Total	146.756 ^a	173.79
Utility Units		
Fuel Gas Production, Cooling, and Treating		26.94
Air Compression		20.47
Steam and Power Generation		30.20
Oxygen Production and Compression		28.93
Raw Water and Cooling Water Systems		12.52
Miscellaneous Utilities		5.67
Total	96.357 ^a	124.73
Off-Site Units		
Ash Dewatering and Storage		6.32
Raw Water Pumping and Storage		4.70
Raw Water Pipeline	12.599	9.51
General Plant Facilities		34.72
Total	60.351 ^a	55.25
Initial Catalysts and Chemicals	2.802	4.01
State Taxes	11.375	- b
Engineering Fees and Licenses	21.595	- b
Start-Up Costs	3.925	4.83 ^c
Contingency at 10%	34.316	36.26
TOTAL DIRECT CAPITAL COST	377.477	398.87
SNG Pipeline to Utility	19.746	- d
Interest During Construction	49.574	46.78
TOTAL	446.797	445.65
Working Capital	9.203	9.13

^aBreakdown not available.

^cAdded to EPNG estimates.

^bIncluded in above.

^dNot included.

Table 4-3. Estimated Annual Operating Expenses and Revenues for the WESCO and EPNG Plants for the Production of SNG from Coal

	Costs, \$ millions/yr		
	WESCO ^a	EPNG	Model Plant ^c
Annual Operating Costs			
Salaries			
Supervision and staff		2.05	
Process operators		3.02	
Maintenance mechanics		4.58	
Materials and Utilities			
River water		0.22	
River pump house power		0.31	
Catalyst and chemicals		1.50	
Feed coal ^d		23.26	
Maintenance materials		7.02	
Supplies		1.05	
Consultant fees		0.21	
O&M for transmission	0.033	-	
Administrative and general	3.355	6.10	
Total O&M	60.946	49.32	55.13
Depreciation, Taxes, Interest and Return on Equity	68.605	80.92	76.15
Total Annual Expense	129.551	130.24	131.28
Revenue			
SNG	120.153	118.55	120.74
By-products	9.398	11.69	10.54
Total	129.551	130.24	131.28
Gas Production Cost, ^b \$/10 ⁶ Btu	1.53	1.51	1.54
, \$/10 ³ scf	1.46	1.44	1.46

^aIncludes transmission over a distance of 67 miles to an existing natural gas pipeline.

^bAssumes production of 78.4×10^{12} Btu/yr of SNG (250×10^6 ft³/day) (330 operating days/yr) (950 Btu/ft³).

^cCapital \$456 million including working capital, 16.7% fixed charge rate on capital.

^dWestern strip-mined coal.

and preparation of environmental impact reports, (3) definitive engineering and construction, and (4) process startup and shakedown. This schedule allows for an estimated "dead time" of 2 years after Title I design for reviews and permits from agencies such as the (1) EPA, (2) Bureau of Mines (coal mining permits), (3) Corps of Engineers (water), Bureau of Indian Affairs (land and land rights), and state governments (building permits, railroads, pipelines, water and air quality, etc.).

A minimum recommended project duration - which might be accomplished if legislation were available to proceed on an emergency basis - is approximately 4.5 years, including (1) 6 months for conceptual design and site selection, (2) 1 year for Title I engineering and environmental reviews, (3) 2.5 years for definitive engineering and construction, and (4) 6 months for plant startup.

5. ALTERNATIVES AND RECOMMENDATIONS

Tables 5-1 and 5-2 present a summary of many of the important physical and economic characteristics of four types of liquefaction processes that are believed to have significant potential for industrial-scale implementation by the early 1980s. Several conclusions have been drawn from these data, together with other considerations expressed in the previous sections:

1. The processes for production of syncrude or boiler fuel by direct catalytic hydrogenation or extraction-hydrogenation appear to have significant potential for commercialization with relatively low technical and economic risk. For each type of process the recommended next scale of development would be a prototype plant with a capacity of 300 to 1000 tons/day of coal. The prototype scale of development could begin immediately for these two types of processes (and perhaps one other, depending on results of further studies) since:

- a. Such projects already have sponsors who will provide a substantial fraction of private funds.
- b. Machinery exists within the government to begin projects of this scale.
- c. The prototype plants constitute a logical next step in scale before pioneer full-scale commercial plants, but are sufficiently small to minimize environmental impacts, permit flexibility in design, and reduce the consequences of technological failure.

2. It appears that the Fischer-Tropsch process for production of gasoline will be too expensive to arouse industrial interest.

3. The data indicate that methanol from coal as a substitute automotive fuel will be less expensive than gasoline from Fischer-Tropsch but probably will be more expensive than gasoline derived from syncrude produced by catalytic hydroliquefaction of coal. The most important advantage of methanol from coal is that the required technology is available now. Thus production on an industrial scale could proceed with a minimum of delay.

Table 5-1. Comparison of Processes Considered for Production of Synthetic Fuels from Coal
All costs expressed in 1973 dollars.

Process or Plant Principal Product Type of Plant Type of Coal Type of Gasifier	Direct Hydrogenation		Extraction/Hydrogenation		Fischer-Tropsch			ICI		WESCO		El Paso
	Synacuda Prototype Bituminous ^a Advanced	Commercial Bituminous ^a Advanced	Boiler Fuel Prototype Bituminous ^a Advanced	Commercial Bituminous ^a Advanced	Commercial Subbit. Lurgi	Commercial Subbit. Advanced	Commercial Subbit. RT	Methanol Commercial Subbit. Lurgi	Commercial Subbit. Lurgi	Commercial Havaho Lurgi	Commercial Havaho Lurgi	
Input												
Is-received coal, tons/day	1,040	35,300	1,144	43,800	148,000	97,500	40,800	25,950	25,650	25,650	25,950	
Electricity, kW	7,450	271,000		400						28,500		
Water, gpm	485	14,300	27	1,000	168,000	111,000	14,800	6,200	5,100	5,100		
Total energy, 10 ⁹ Btu/day	23.2	782			2,600	1,710	724	461	440	440		
Output												
High Btu fuel gas, 10 ⁹ Btu/day		110	6.2	101,520	100,000	100,000	154,830	38,700				
Liquid fuel, bbl/day	2,260	82,310	2,100	18,520	12,600	12,600	20,000	5,000				
By-products, tons/day	13.8	470	11.5	627	574	574	400	2,200	1,580			
Waste gases, tons/day	18	1,366		1,280	8,600	8,600			17,400			
Waste water, gpm	130	4,900	100	77,720	30,700	20,300	12,000	1,000	1,000			
Solid wastes, tons/day	120	4,100	150	4,260	28,100	18,500	2,000	4,700	6,400			
Total Capital Investment, \$ millions	60.5	685	62	832	2,293	1,198	722	521	456	455		
Annual Operating Cost, \$ millions ^c	13.3	3.38	30.5	4.02	12.1	11.91	5.83	5.78	5.69	5.80		
Fuel Production Costs												
High-Btu gas, \$/10 ⁶ Btu	5.3	1.57	3.70	1.79	4.58	3.07	2.05	1.87				
Liquid fuel, \$/10 ⁶ Btu	32.1	8.94	20.3	11.0	24.2	16.2	5.60	5.08				
Process Efficiency, %	60	74	65 ^e	63	22.1	33.5	45	59	67	63		
Manpower	96	930	96	1100								
Engineering, m/y	390	3,710	390	4,000							635 ^d	
Operation, m/y/yr	280	2,800	300	3,000							5,700	
Steel for Construction, 10 ³ tons	10	96	11	130							50-60	

^aMidwestern Bituminous Coal at \$7.50/ton.

^bWestern strip-mined subbituminous coal at \$3.00/ton.

^cIncluding by-product credits.

^dLine requires 36 m/y for engineering, 1500 m/y for construction, and 40,000 tons steel.

Table 5-2. Estimated Effects of Type of Process, Type of Coal, Cost of Coal, and Capital Fixed Charge Rate on the Cost of Producing Several Synthetic Fuels from Coal

Product	Process	Coal		Capital Fixed Charge Rate %	Product Cost	
		Type	Cost \$/Ton		\$/10 ⁶ Btu	\$/
Synocrude ^a	H-Coal	Bituminous	7.50	23.4	1.57	
		Bituminous	8.50	23.4	1.63	
		Bituminous	7.50	17.9	1.37	
		Subbituminous	3.00	23.4	1.59	
		Subbituminous	4.00	23.4	1.68	
		Subbituminous	3.00	17.9	1.36	
Boiler Fuel	SRC, Consol	Bituminous	7.50	23.4	1.79	
		Bituminous	8.50	23.4	1.86	
		Bituminous	7.50	17.9	1.56	
Gasoline	FT-Lurgi	Subbituminous	3.00	23.4	4.58	
		Subbituminous	4.00	23.4	4.87	
		Subbituminous	3.00	17.9	3.85	
	FT-Bi Gas	Subbituminous	3.00	23.4	3.07	
		Subbituminous	4.00	23.4	3.26	
		Subbituminous	3.00	17.9	2.60	
Methanol	KT-ICI	Subbituminous	3.00	23.4	2.06	
		Subbituminous	4.00	23.4	2.16	
		Subbituminous	3.00	17.9	1.74	
	Lurgi-SNG	Subbituminous	3.00	23.4	1.87	
		Subbituminous	4.00	23.4	1.97	
		Subbituminous	3.00	17.9	1.67	
SNG	Lurgi			23.4	1.93	
				17.9	1.61	
				16.7	1.54	

^a Gasoline produced from this material would be higher in cost by about \$2.0 to 2.5/bbl.

Recommended Demonstration Program. - The recommended schedules and incremental funding for a joint government-industry program to demonstrate liquefaction technology are presented in Tables 5-3 and 5-4. Specific objectives of this program are as follows:

1. Conduct joint programs from July 1974 to July 1977 to design, construct, and shakedown three prototype (capacity 300 to 1000 tons coal/day) plants - H-Coal Syncrude, SRC-Hydrogenation for boiler fuel, and a third type (a modified CSF or hydrocarbonization). These plants are believed to represent the optimum size to provide the fastest and minimum risk path from present development work to commercial plants. Such prototype plants are generally favored by industry and substantial (perhaps 1/3) investment of private funds can be expected because of the relatively low (about \$60 million) capital costs of these plants.

The present plan is that these prototype plants would be built at sites that are already highly industrialized (petroleum refineries, petrochemical complexes, or large power plants) and, thus, will have small incremental environmental impact. These sites will have (a) trained personnel, (b) technical services, (c) supplies of coal and/or other raw materials, and (d) facilities for testing of the products.

2. Conceptually design, collect base line data on sites, and prepare generic environmental impact statements for two pioneer hydroliquefaction plants (~50,000 tons coal/day) in the period July 1974 to July 1976. Design, construct, and shakedown these two pioneer plants in the period July 1976 to July 1979. Government funding of about 20% of the capital costs of these plants may be required because there is some risk that engineering funds would be wasted since the engineering would begin before operation of the prototype plants. It may also be necessary to waive the formal environmental impact review for these plants.

3. After detailed conceptual design and siting studies, engineer, construct, and shakedown a pioneer commercial (~5000 tons MeOH/day) plant in the period January 1975 to January 1979.

Detailed recommendations on the production of methanol from coal have been formulated and are contained in a separate report.

Table 5-3. Incremental Government Funding for the Recommended Program for Demonstrating the Production of Synthetic Fuels from Coal.

	Fiscal Year Costs, \$Millions					Total
	1975	1976	1977	1978	1979	
<u>ER & D Program for Coal</u>						
Coal Liquefaction	75	75	75	75	75	375
R&D, Other Operating Expenses	(57)	(45)	(43)	(46)	(55)	(246)
Construction-SRC Pilot Plant	(4)	-	-	-	-	(4)
Direct Hydrogenation						
Prototype Plant	(8)	(20)	(18)	(3)	-	(49)
Advanced Process Prototype Plant		(2)	(8)	(23)	(20)	(53)
Multiple Process Pilot Plant	(6)	(8)	(6)	(3)	-	(23)
Two Synthetic Fuel Pioneer Plants	100	100	55	50	50	355
High Btu Gasification	35	75	92	81	57	340
R&D, Other Operating Expenses	(14)	(25)	(47)	(49)	(53)	(188)
Construction-Hygas Pilot Plant	(2)	(2)	(4)	(2)	-	(11)
- CO ₂ Acceptor Pilot Plant	(2)	-	-	-	-	(2)
- Synthane Pilot Plant	(8)	(7)	(2)	-	-	(17)
- Bi-Gas Pilot Plant	(9)	(12)	-	-	-	(21)
- Demonstration Pilot Plant		(28)	(39)	(30)	(4)	(101)
Mining	45	57	64	77	82	325
Direct Combustion	30	35	40	44	51	200
Low Btu Gasification	30	37	42	48	43	200
Environmental Central Technology	70	50	42	45	53	260
Supporting R&D	20	22	24	27	27	120
	405	451	434	447	438	2175
<u>Recommended Incremental Funding for Synthetic Fuels Demonstration Program</u>						
Construction ^{a/}						
Direct Hydrogenation Prototype Plant	2	5	-13	-3		-9
Extraction/Hydrogenation Prototype Plant	10	25	5			40
Third Prototype Plant	10	23	-3	-23	-20	-13
Methanol Pioneer Plant	0	0	0	0	0	0
Second Pioneer Plant	0	0	150	0	0	150
Third Pioneer Plant	0	0	150	0	0	150
Fischer-Tropsch Design Studies	5	15	-	-	-	20
First Generation SNG Production	5	20	20	10	10	65
Supporting R&D, Operations	15	20	27	35	25	120
	47	55	336	19	15	525

^{a/} Assumes government provides 2/3 and 20% of capital for prototype and pioneer plants, respectively.

4. Conceptually design and perform Title I engineering of a commercial Fischer-Tropsch gasoline plant in the period July 1974 to July 1976 for reference and standby construction.

5. Accelerate the pace of supporting research and development of liquefaction as a coordinated extension of the \$10 billion energy R&D program.

The recommended synthetic fuels development program also includes provisions for accelerating the development of the industry for the production of SNG from coal. The recommended activities include (a) engineering development of a slagging Lurgi gasifier, (b) engineering development of a high-pressure Koppers-Totzek gasifier (to permit wider use of coals), (c) site studies (collecting base line data on potential commercial sites), (d) preparation of generic environmental impact statements, (e) accelerated research in environmental effects, (f) development of advanced materials, catalysts, and equipment components, (g) additional development of the methanation step, (h) institutional research, (i) development of services and capabilities for process vessel manufacturing, and (j) development of standards to minimize the routine design requirements.