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ASSESSMENT OF H-COAL PROCESS DEVELOPMENTS

Impact on the Performance and Economics of a
Proposed Commercial Plant

By
Abu Talib
David Gray
Martin Neuworth

January 1984

Work Performed Under Contract No. AC01-80ET13800

The MITRE Corporation
Metrek Division
McLean, Virginia

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ABSTRACT

This report assesses the performance of the H-Coal process, a catalytic direct liquefaction process, at a process development and large pilot-plant scale of operation. The assessment focused on the evaluation of operating results from selected long-term successful process development unit (PDU) and pilot plant runs made on Illinois No. 6 coal. The pilot plant has largely duplicated the product yield structure obtained during the PDU runs. Also, the quality of products, particularly liquid products, produced during the pilot plant run is quite comparable to that produced during the PDU runs. This confirms the scalability of the H-Coal ebullated-bed reactor system from a PDU-scale, 3 tons of coal per day, to a large pilot scale, 220 tons of coal per day, plant.

The minor product yield differences, such as higher yields of C₃, C₄, and naphtha fractions, and lower yields of distillate oils obtained during pilot plant runs as compared to the PDU runs, will not impact the projected technical and economic performance of a first-of-a-kind commercial H-Coal plant. Thus, the process yield and operating data collected during the PDU operations provided an adequate basis for projecting the technical and economic performance of the proposed H-Coal commercial plant.

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EXECUTIVE SUMMARY

The H-Coal process, a catalytic direct-liquefaction process, is being developed by Hydrocarbon Research, Inc. (HRI), as an outgrowth of their H-Oil process, a commercially demonstrated technique for upgrading petroleum residuum. The H-Coal technology has undergone extensive developmental efforts from small, bench-scale tests to process development unit (PDU), and finally large-scale, pilot plant operations. Preliminary process design and economics have also been developed under a Phase 0 effort for a proposed first-of-kind, H-Coal commercial plant. Since the commercial plant design was developed prior to the pilot plant operations, the operating results of a selected PDU run were used to project the technical and economic performance of the proposed commercial plant. Since that time, the H-Coal process has been demonstrated on a 200-TPD pilot plant scale at Catlettsburg, Kentucky. Process operating data from the pilot plant are now available as input to the Phase 0 commercial plant design.

The U.S. Department of Energy (DOE) requested MITRE to carry out an independent assessment to evaluate the performance of the H-Coal process at the large-scale pilot plant, and its subsequent impacts on the configuration, design, operation, and projected economics of the proposed commercial plant.

Process Description

The key unit operation of an H-Coal direct-liquefaction process is the ebullated-bed reactor system which facilitates the catalytic conversion of coal to liquid products. A simplified schematic diagram of the H-Coal process is given in Figure S-1. The coal feed is properly sized, dried, and mixed with recycle slurry and process-derived solvent (normally a part of the heavy distillate oil product). The coal/oil slurry, along with part of the recycle hydrogen, is preheated to initiate the coal dissolution, and the preheated slurry is introduced in the bottom of the ebullated-bed reactor. The remaining hydrogen, after being preheated in a separate heater, is also introduced through the bottom of the reactor.

In the ebullated-bed reactor, the coal feed, in the presence of a cobalt or nickel and molybdenum catalyst, is further hydrocracked to gaseous, light, intermediate, and heavy liquid products. The ebullated-bed mechanism keeps the catalyst in a state of constant fluidization, and enables the on-line withdrawal and addition of fresh catalyst. Thus, an overall uniform catalytic activity is maintained within the reactor, which is crucial for the effective conversion of coal to distillate liquid products. The ebullated bed also promotes intimate contact of catalyst, hydrogen, and coal solution, and maintains nearly isothermal conditions within the reactor.

Source: Reference 3

FIGURE S-1
SCHEMATIC OF THE H-COAL PROCESS
FOR THE SYNCRUDE MODE

The reactor products, including the unconverted coal and mineral matter, are processed in a series of letdown vessels that operate at progressively lower pressures. The separated gaseous and liquid products, both light and heavy, are further processed to meet the commercial and/or in-plant recycle specifications. The coal/oil slurry from the low-pressure letdown vessel is partially concentrated in a hydroclone system; the hydroclone overflow (HO) stream, which contains part of the unconverted coal and ash, along with a portion of heavy oil product, is used to slurry the fresh coal feed. The hydroclone underflow is vacuum flashed to produce bottoms containing primarily unreacted coal (IOM), 975⁺°F distillate, and ash. This material is gasified to provide the hydrogen requirement of the plant.

H-Coal Process Development Efforts

The H-Coal process development efforts began in 1965, gradually progressing through bench-scale testing, process development stages, and successful pilot-scale operations to the preliminary design of a proposed first-of-kind commercial plant. Over twenty different coals have been tested in more than 60,000 hours of combined operation of the bench-scale and PDU units. The bench-scale, coal-liquefaction unit handles 25 pounds of coal per day and is primarily used for process improvement studies, including catalyst evaluation and testing new coals. The 3.5-TPD capacity PDU has been used primarily for larger, continuous test runs to develop design basis and operating parameters for the large Catlettsburg pilot plant. Also, the same design basis was later used to develop a preliminary design, or Phase 0 design, for the commercial plant. Specifically, a successful longer term, about 30 days, PDU operation (PDU Run 5) on Illinois No. 6 coal, provided the basis for both the pilot plant and Phase 0 commercial plant designs.

The major objectives of the Catlettsburg pilot plant were to provide process and mechanical data for the final design construction and operation of a first H-Coal commercial plant. The pilot plant was commissioned in March, 1980. Operation on coal commenced in May, 1980. During the two and one-half years of operation between May, 1980 and November, 1982, the pilot plant has logged approximately 7,000 hours of successful operation, while processing three coals (Illinois No. 6, Kentucky 9/11, and Wyodak). The longest pilot plant run (Run 8) lasted 131 days, during which time an average of 214 tons per day of coal, mostly Illinois No. 6, were processed. During Run 8, steady-state operating conditions were sustained and a material balance closure of about 98 percent was obtained. Run 8 demonstrated the mechanical operability and reliability of critical equipment.

The first H-Coal commercial plant was proposed to be located in Breckinridge County, Kentucky, and H-Coal commercialization efforts were carried out under the Breckinridge project. As part of the

Phase 0 effort, preliminary design and economics were developed for the proposed Breckinridge project. As indicated earlier, PDU operating results were used to develop the preliminary design under Phase 0 efforts. Several attempts to obtain financial support, i.e., loan guarantee, for the Breckinridge project from the Synthetic Fuels Corporation were unsuccessful because of insufficient equity participation. The Breckinridge project, for all practical purposes, has been indefinitely postponed.

Comparison of PDU and Pilot Plant Operations

The H-Coal process development unit (PDU) is used for intermediate-scale process development efforts. The heart of the PDU is an ebullated-bed reactor which is used for the catalytic hydro-conversion of slurried coal to distillate quality liquid products. The PDU contains all the required unit operations, i.e., coal feed preparation, slurry preparation, reactor product separation, etc. The ebullated-bed reactor system, along with its supporting unit operations, are duplicated in the larger capacity pilot plant.

In order to assess the performance of the H-Coal process at two scales of operation, PDU Run 5 and Pilot Plant Run 8 are compared. These two runs were primarily chosen because PDU Run 5 provided the basis for the commercial Phase 0 Breckinridge design and Pilot Plant Run 8 represented the longest period of operation using Illinois No. 6 coal. These selected runs achieved smooth plant operations with sustained, steady-state product yields. Specific periods of operation during PDU Run 5 and Pilot Plant Run 8 were selected to carry out the H-Coal process performance comparison. For PDU Run 5, Period 29 (i.e., the 29th day from the start of the run) was selected to represent conditions close to equilibrium with respect to catalyst activity and product yields. The special four-day material balance period of Pilot Plant Run 8 was selected to represent comparable equilibrium conditions. The operating conditions during these periods of selected plant runs are compared in Table S-1. Both the selected runs were conducted on Illinois No. 6 coal.

The operating conditions during the specific periods of operation are quite comparable, except for the reactor operating pressure and cumulative catalyst age. Even though the reactor operating pressure is higher for the pilot plant than for the PDU, the hydrogen partial pressure is comparable. A higher cumulative catalyst age was obtained for the pilot run when compared to the PDU run, thus indicating that the pilot run was closer to the catalyst equilibrium activity.

TABLE S-1

COMPARISON OF OPERATING CONDITIONS FOR SELECTED
PDU AND PILOT PLANT RUNS

<u>Run Designation</u>	<u>PDU Run 5, Period 29</u>	<u>Pilot Plant Run 8, 4-Day Mat'l Bal. Period</u>
	October 2-3, 1977	Oct. 29-Nov. 2, 1981
Coal Type	Illinois No. 6	Illinois No. 6
Average Reactor Temp., °F	851	848
Reactor Pressure, PSIG	2,700	3,006
Hydrogen Partial Pressure, PSIG		
o Inlet	2,490	2,512
o Outlet	1,920	2,058
Cumulative Catalyst Replacement Rate, lb/ton dry coal	1.06	0.96
Cumulative Catalyst Age, lb coal/lb catalyst	1,102	1,695
Coal Space Velocity, lb/hr/ft ³	31.46	29.2
Recycle Slurry Composition		
o Hydroclone Overflow	65%	58%
o Atmospheric Still Bottoms	35%	42%

The operating results obtained during the two specific periods of operation are compared in Table S-2.* The overall distillate (C_4^+ - 975°F) yield obtained during the PDU and pilot runs are in excellent agreement. The C_4 - 975°F distillate oil yields (on a weight basis) for both runs were higher than 47 percent of the dry coal fed to the H-Coal reactor. The volumetric yield of the C_4^+ - distillate material is about 5 percent higher for the pilot plant than for the PDU. The pilot plant, however, produced approximately 81 percent and 34 percent more C_4 and C_3 fractions, respectively, than did the PDU. Because of the higher C_3 and C_4 fraction yields, the overall hydrogen consumption has increased from 4.9 percent of the dry coal input for the PDU run to 5.47 percent of the dry coal for the pilot run, an increase of about 11 percent.

The elemental composition and other physical properties for different distillate fractions produced during the PDU and pilot runs are comparable. The pilot plant distillate products, particularly naphtha (180°F to 380°F) and light distillate oil (380°F to 650°F) are slightly lighter than the comparable PDU products. Also, the PDU products contained more nitrogen and less oxygen than the comparable pilot plant products. These minor differences in the elemental composition of the liquid products may be a direct result of the differences in the elemental composition of feedstock coals.

H-Coal Commercial Plant Design and Cost Basis

The proposed H-Coal commercial plant is a fully-integrated, grassroots facility. This includes provisions for on-site beneficiation and handling of run-of-mine (ROM) coal, conversion of coal into liquid and gaseous hydrocarbons, separation and purification of the products, stabilization and reforming of the naphtha, recovery of byproduct sulfur, ammonia, and phenols, production of process hydrogen by gasification of vacuum bottoms, and plant steam and utility production by on-site, coal-fired boilers. Additionally, wastewater treatment facilities, air emission controls, and solid waste disposal are integral parts of the process design.

A simplified block flow diagram for the Phase 0 design is given in Figure S-2. As shown in this figure, some of the blocks represent several plants and support facilities. The Breckinridge Phase 0 commercial plant is designed to process approximately 23,000 tons per day of run-of-mine coal, to produce approximately 46,000 barrels per day of C_4^+ -distillate products. Major products and byproducts from the proposed plant include pipeline quality gas (SNG), propane, butane, light straight-run (LSR) gasoline, reformate, distillate oils, ammonia, and sulfur.

*For comparison, the product distribution obtained during Pilot Plant Run 11 is also shown.

TABLE S-2

COMPARISON OF OPERATING RESULTS FROM
SELECTED H-COAL PDU AND PILOT RUNS

PLANT OUTPUT	Run Designation				
	PDU		PILOT PLANT		
	Run 5, Period 29		Run 8, 4-day Material Bal.		Run 11*
	LBS/100 LBS OF DRY COAL	BBL/TON OF DRY COAL	LBS/100 LBS OF DRY COAL	BBL/TON OF DRY COAL	LBS/100 LBS OF DRY COAL
H ₂	-4.91		-5.47		-4.74
H ₂ S	2.64		2.67		2.63
NH ₃	1.08		0.65		1.08
H ₂ O	6.67		6.67		6.67
CO ₂	0.48		--		0.48
C ₁	3.75		4.20		3.75
C ₂	3.54		3.02		2.94
C ₃	3.39		4.55		3.20
C ₄	2.54		4.61		1.88
C ₅ - 400°F	16.20	1.40	17.8	1.69	16.67
400- 650°F	20.37	1.21	16.46	0.99	19.30
650- 975°F	7.96	0.41	8.81	0.46	10.39
Residuum	19.00		21.26		21.68
Unconverted Coal	5.78		3.46		3.51
Ash	11.51		11.31		10.58
TOTAL:	100.00		100.00		100.00
Total Distillate, C ₄ ⁺ - 975°F	47.1	3.01	47.7	3.14	48.24
Residuum, Unconverted Coal and Ash	36.3		36.0		35.8

*Mr. Richard Santore of PETC, in reviewing this manuscript, called attention to the results of Pilot Plant Run No. 11. The results of this run agree more closely with PDU Run No. 5 in terms of hydrogen consumption. The conclusions of this report with respect to the negligible impact of deviations in Run No. 8 from PDU No. 5, are still unaltered.

FIGURE S-2
OVERALL CONFIGURATION FOR THE PROPOSED
PHASE 0 H-COAL COMMERCIAL PLANT

The overall material balance and flow quantities of products and byproducts for the Phase 0 design were based on the overall product yield structure obtained during Period 29 of PDU Run 5. The overall products/byproducts output for the Phase 0 plant is given in Table S-3. The total capital cost summary for the Phase 0 plant is given in Table S-4. The capital cost data are extracted from the detailed cost estimates developed by Bechtel, Inc., under the Breckinridge project initial effort. The total capital requirements for the Phase 0, H-Coal commercial plant are presented both in constant 1981 dollars and in "as spent", or current dollars. The constant 1981 dollars represent the overnight construction costs, which do not take into account the actual capital expenditures over the design/engineering, procurement, construction, commissioning, and startup periods.

TABLE S-3

OVERALL PRODUCT/BY-PRODUCT OUTPUT
FOR THE PHASE 0 COMMERCIAL PLANT

<u>Coal Input to the Liquefaction Section</u>		<u>16,500 Tpsd</u>
Output:		
● Propane		6,524 Bpsd*
● Mixed Butane		4,303 Bpsd
● Light Naphtha		3,742 Bpsd
● Reformate		11,421 Bpsd
● Distillate Oil		26,359 Bpsd
● By-products		
	SNG	45.8 MMscfd**
	Sulfur	534 LTpsd
	Ammonia	193 Tpsd

*Barrels per stream day.

**Includes refinery-rich gas

TABLE S-4

BASE CASE TOTAL CAPITAL REQUIREMENTS (1981-1988)

	<u>1981 Dollars</u>	MM\$	<u>Current Dollars</u>
Equipment and Construction	3,167.00		4,491.00
Working Capital	91.00		36.00**
Interest during Construction	<u>--</u>		<u>1,407.00</u>
TOTAL	3,258.00*		5,934.00***

*Overnight construction costs

**Working capital required in 1988

***Overall funding requirement

The total capital requirement for the Phase 0 plant is 3,258 million dollars (January 1981 dollars). The total "as spent" cumulative capital requirement is about six billion dollars.

The annual operating costs, as estimated by Bechtel, Inc., are presented in Table S-5. The total annual estimated operating costs are about \$452 million dollars.

TABLE S-5

ANNUAL OPERATING COST

(87% Stream Factor - January 1981 Dollars))

I. Coal Feedstock and Utilities Costs*	323,615,100
II. Labor Costs	39,574,300
III. Supplies and Materials	32,254,200
IV. Other Costs	
o Indirects** and G&A	8,782,500
o Local Taxes and Insurance	47,511,500
TOTAL Annual Operating Cost	<u>451,737,600</u>

*As received run-of-mine coal.

**Indirect costs include items such as laboratory, safety, health, and security, etc. related expenses.

Impacts of Pilot Plant Operations on Phase 0 Commercial Plant Design and Economics

The operating results obtained during Pilot Run 8, the four-day material balance period, are incorporated in the Phase 0, H-Coal commercial plant design basis. The overall configuration of the Phase 0, H-Coal commercial plant described earlier will not change as a result of incorporating the pilot plant operating results. However, some design modifications will be required to incorporate the pilot plant operating results. The significant modifications are:

- Revision of overall product yield structure
- Expansion of gas plant to accommodate increased propane and butane production
- Expansion of gasifier plant to process slightly higher quantities of vacuum bottoms
- Small increase in oxygen plant capacity
- Expansion of steam plant to meet the increased high-pressure steam demand of the oxygen plant.

A comparison of the overall products/byproducts output expected from the modified Phase 0 plant compared to the Phase 0 design is given in Table S-6. The propane and butane production rate for the modified Phase 0 plant has increased by 1,722 and 3,569 Bbl/day, respectively. The reformat production has increased by 1,161 Bbl/day. The distillate oil production, however, has decreased by about 5,500 Bbl/day.

Because of the required capacity expansions in the gas plant, the gasification or hydrogen plant, and the oxygen plant, the utility consumption of these plants will proportionately increase. The impacts on the utilities demand and thermal efficiency of the Phase 0 plant are given in Table S-7. The steam consumption of the modified plant has risen by about 8 percent over the Phase 0 plant requirement. Approximately one percent more run-of-mine coal will be required to fuel the boiler plant. The purchased electric power consumption has increased by about 6 percent over the Phase 0 consumption. The overall thermal efficiency, a ratio of the higher heating value of products, including SNG, divided by the higher heating value of coal and the electrical energy input, has decreased from 61 percent for the Phase 0 plant to 58 percent for the modified Phase 0 plant. The decrease in thermal efficiency is a direct result of increased utility consumption and decreased production of higher energy content, distillate oil product for the modified plant. However, while the overall thermal efficiency has decreased for the modified plant, the yield of readily marketable

TABLE S-6

OVERALL PRODUCT/BY-PRODUCT OUTPUT
FOR THE PHASE O AND MODIFIED PHASE C COMMERCIAL PLANT

	<u>Phase O Commercial Plant</u>	<u>Modified Phase O Commercial Plant</u>
Coal Input to the Liquefaction Section		16,500 Tpsd
Output:		
Propane	6,524 Bpsd*	8,246 Bpsd
Mixed Butane	4,303 Bpsd	7,871 Bpsd
Light Naphtha	3,742 Bpsd	2,831 Bpsd
Reformate	11,421 Bpsd	12,561 Bpsd
Distillate Oil	26,359 Bpsd	20,585 Bpsd
By-Products		
SNG	45.8 MMscfd**	45.8 MMscfd**
Sulfur	534 LTpsd	534 LTpsd
Ammonia	193 Tpsd	193 Tpsd

*Barrels per standard day.

**Includes refinery-rich gas.

liquid products, i.e., propane and reformat, has significantly increased. Therefore, the thermal efficiency loss will not significantly penalize the overall performance of the modified commercial plant.

The hydrogen utilization efficiency (percent of total input hydrogen used in the production of C_4^+ -liquid products) rather than the overall thermal efficiency, is a better indicator of the performance of a direct liquefaction process. The hydrogen utilization efficiency for both Phase 0 and modified plants is compared in Table S-8. For both the Phase 0 and modified plants, approximately 50 percent of the total input hydrogen to the liquefaction plant is used to produce C_4^+ -distillate liquid products. Therefore, the hydrogen utilization efficiency of the commercial plant has not changed because of the required plant modifications. On the other hand, the amount of hydrogen used for producing C_1 - C_3 hydrocarbons has increased from 23 percent of the total input for the Phase 0 plant to about 24 percent for the modified Phase 0 plant.

The economic impacts of the required Phase 0 design modifications are also estimated. The economic performance of the Phase 0 commercial plant was projected by Bechtel, Inc., under the Breckinridge project initial effort. MITRE has used the results of Bechtel's economic analysis to quantify the economic impacts, if any, resulting from the Phase 0 design modifications discussed earlier.

The impacts of Phase 0 design modifications on the capital and operating cost of the commercial plant are given in Table S-9. The Phase 0 plant modifications have increased the total capital and operating costs by approximately two and one percent, respectively. The assumptions used by Bechtel, Inc. for the Phase 0 plant economic analysis are given in Table S-10. The impacts of Phase 0 plant modifications on the projected economic performance of the commercial plant are shown in Table S-11.

The Phase 0 plant modifications have decreased the annual revenues by approximately one percent. The corresponding reductions in the discounted cash flow rate of return on equity (DCF-ROE) are insignificant. For example, for 100-percent equity financing, the DCF-ROE has decreased from 12.2 percent for the Phase 0 plant to 12.0% for the modified Phase 0 plant. Similarly, for the assumed debt financing, the DCF-ROE has decreased from 20.8 percent for the Phase 0 plant to 20.2 percent for the modified Phase 0 plant.

Conclusions

The scale-up of the H-Coal process has been demonstrated in the large pilot plant. The scale-up factors used for the smaller PDU and large-scale pilot plant, along with the expected scale-up for a single

TABLE S-7

IMPACTS ON UTILITIES DEMAND AND THERMAL EFFICIENCY

	<u>Phase 0 Design</u>	<u>Modified Phase 0 Design</u>	<u>Percent Change</u>
Total Plant Coal Input	23,582 Tpd (D.B.) (ROM)	23,733 (D.B.) (ROM)	+0.6
Total High Pressure Steam Production (900 psig, 750°F)	1.2×10^6 lbs/hr	1.3×10^6 lbs/hr	+8.3
Total Electric Power Consumption	170,000 kWh/hr	179,600 kWh/hr	+6
Overall Thermal Efficiency	61%	58%	-3

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TABLE S-8

HYDROGEN UTILIZATION EFFICIENCY*
FOR PHASE 0 AND MODIFIED PHASE 0 PLANTS

	<u>Phase 0 Plant</u>		<u>Modified Phase 0 Plant</u>	
	<u>lbs of H₂/</u> <u>100 lbs of Coal</u>	<u>Percent</u>	<u>lbs of H₂/</u> <u>100 lbs of Coal</u>	<u>Percent</u>
<u>Input:</u>				
Coal Feed	4.96	50.2	4.96	47.5
H ₂ Consumed in the Process	<u>4.91</u>	<u>49.8</u>	<u>5.47</u>	<u>52.5</u>
TOTAL H ₂ INPUT	9.87	100.0	10.43	100.0
<u>Output:</u>				
C ₁ - C ₃	2.23	22.7	2.49	23.8
C ₄ - Liquid Products	4.97	50.2	5.26	50.6
Vacuum Bottoms	1.44	14.6	1.44	13.8
H ₂ , NH ₃ , H ₂ S, Fuel Gas and Losses	<u>1.23</u>	<u>12.5</u>	<u>1.24</u>	<u>11.8</u>
TOTAL OUTPUT	9.87	100.0	10.43	100.0

*Percent of input hydrogen in the C₄⁺ - liquid products.

TABLE S-9
TOTAL CAPITAL REQUIREMENT AND ANNUAL OPERATING COSTS
(First Quarter 1981 Dollars)

	Phase 0 Design (MM \$)	Modified Phase 0 Design (MM \$)	Percent Change
Total Capital Requirement	3,258.00*	3,326.00**	+2.0
Total Annual Operating Costs	451.7*	456.5**	+1.0

*Bechtel, Inc. estimate.

**Add-on cost for the modified plant as estimated by MITRE.

TABLE S-10
BASE CASE ECONOMIC ASSUMPTIONS*

Products Price Forecast: Based on crude oil valued at \$35/Bbl (1981\$)
increasing to \$41.50/Bbl by 1990, and remaining at
that value through the life of the project

Inflation: Average 8.5%/yr through construction, and 6.0% during
operating life of the project

Tax Credits: 15% of total fixed capital

Tax Depreciation: Project life, 15 years; sum of years digits followed by
double declining balance

Plant Stream Factors: 20% - 1988, 50% - 1989, 80% - 1990, 86.4% - 1991,
through project life

Type of Financing: 100% equity
48/52% Debt/Equity**

Interest Rate on Debt: 10.6%

*Used by Bechtel, Inc.

**Corresponds to a \$3 billion loan limit set by the U.S. Synthetic Fuels
Corporation (SFC)

TABLE S-11

ECONOMIC IMPACTS OF PHASE ZERO PLANT MODIFICATIONS
RESULTING FROM PILOT PLANT OPERATIONS*

(Current Dollars)

<u>Type of Financing</u>	<u>Phase 0 Plant</u>		<u>Modified Phase 0 Plant</u>	
	<u>Total Revenue**</u>	<u>DCF-ROE</u>	<u>Total Revenue**</u>	<u>DCF-ROE</u>
100-percent Equity	\$1,870* MM	12.2%	\$1,860 MM	12.0%
Debt Financing to \$3 billion SFC Loan Limit	\$1,870 MM	20.8%	\$1,860 MM	20.2%

*Results for Phase 0 plant are taken from Blue Books, whereas the results for modified plant were estimated by MITRE, using Bechtel's economic analysis.

**Composite price for C₃ - 850 liquid products for Phase 0 plant is \$94.4/Bbl in 1991.

Composite price for C₃ - 850 liquid products for Modified Phase 0 plant is \$93.7/Bbl in 1981.

xxxx
1

commercial-scale reactor, are given in Table S-12. The successful operation of the large-scale, H-Coal pilot plant demonstrates scale-up factors of 7 and 73, based on reactor diameter and throughput, respectively. The corresponding scale-up factors between the pilot plant and the proposed pioneer commercial plant are 2.3 and 11, respectively. Therefore, the H-Coal commercial plant is more likely to duplicate the process performance achieved during the pilot plant operations, thus reducing the technical risk associated with the proposed first-of-a-kind project.

Even though the technical risk is minimized, the economic risk associated with the proposed project is still quite significant. The implementation of the project requires approximately six billion dollars. With 100-percent equity financing, the project is not economically viable. The assumed debt financing, however, improves the economic viability of the project significantly.

TABLE S-12

H-COAL REACTOR SCALE-UP FACTORS

	H-Coal Reactor Diameter	Throughput (Coal)	Scale-up Factors	
			Based on Diameter	Based on Throughput
PDU	8.5 in.	3 TPD	10.4*	240*
Pilot Plant	5 ft.	220 TPD	7.0	73.0
Commercial Plant	10-13 ft.	2,357 TPD**	2.3	11

*Represents approximate scale-up between bench-scale and PDU unit.

**Capacity for each of seven on-line reactors.

The differences in the operating results of the pilot plant and the PDU were not significant enough to change the technical and economic performance of the Phase 0 commercial plant as developed under the Breckinridge project initial effort.

Recommendations

Even though the H-Coal process is ready for commercial implementation, there are opportunities for selected research and development (R&D) efforts to improve the overall liquid yield and process economics. These R&D opportunities and recommendations for future work are given below.

H-Coal Process Improvements

R&D Opportunities:

- The extinction of 650–850°F distillate oil fraction, vacuum gas oil (VGO), by selective recycle to the H-Coal reactor offers potential for increasing the selectivity to C₄–650°F liquid product. The C₄–650°F syncrude can subsequently be upgraded to gasoline and diesel fuel by a combination of high-severity hydrotreating and reforming unit.

R&D Recommendations:

The operability of the H-Coal process in the VGO extinction mode should be tested both at bench- and PDU-scale operations so that meaningful performance data, i.e., reactor space velocity, temperature, pressure, catalyst replacement rate, etc., can be obtained to confirm the potential advantages offered by this scheme.

General Liquefaction Studies

R&D Opportunities:

- In the present configuration, the H-Coal process uses only 50 percent of the total input hydrogen for producing C₄⁺-distillate product. Approximately 23 percent of input hydrogen is consumed in the C₁–C₃ hydrocarbon production.
- A significant reduction in C₁–C₃-make would increase liquid fuel yields, reducing overall hydrogen consumption and possibly product costs.
- Performing the diverse thermal and catalytic reactions in a single reactor subjected to fixed reaction conditions is primarily responsible for high C₁–C₃-make.
- This strongly suggests an advantage in carrying out the overall process in more than one stage.

R&D Recommendations:

- Continued basic research studies are recommended in order to optimize the various steps involved in the transition of coal to clean distillate fuel.
- These studies should be coordinated with multistage liquefaction process developments currently being studied by Lummus, Chevron, Kerr McGee, and HRI.

1.0 INTRODUCTION

The H-Coal liquefaction process is being developed by Hydrocarbon Research, Inc. (HRI), as an outgrowth of their H-Oil process, a commercially demonstrated technique for upgrading petroleum residuum. The H-Coal technology has been under continuous development since the early bench-scale tests conducted some twenty years ago. The objectives of these developmental efforts were to provide equipment design and operational data required for the successful design, construction, and operation of a commercial-scale H-Coal liquefaction plant.

The H-Coal liquefaction process is at the present time a candidate for commercial demonstration. The first H-Coal commercial plant is proposed to be located in Breckinridge County, Kentucky. The H-Coal commercialization efforts are being carried out under the Breckinridge project. The proposed commercial H-Coal plant is designed to process approximately 23,000 tons per day of run-of-mine (ROM) coal, to produce a nominal 50,000 barrels per day of liquid products. A preliminary engineering design, i.e., a Phase 0 design, has been developed for the Breckinridge project.⁽¹⁾ Product yields and distribution used as the basis for the original Phase 0 design were obtained from a process development unit (PDU) processing three tons of coal per day on the candidate coal feedstocks. Since that time, the H-Coal process has been demonstrated on a 200-TPD pilot plant scale at Catlettsburg,

Kentucky. Process operating data at this scale are now available as input to the projected commercial-scale performance used in the Phase 0 design.

The purpose of this study is to evaluate the impact of process yield and operating data collected during the pilot plant operations, on the projected performance and economics of the Phase 0 commercial design. The U. S. Department of Energy (DOE) requested MITRE to carry out this study as a lessons-learned assessment to guide future DOE-supported research and development efforts. The scope of the study includes:

- Review of operating data obtained in selected PDU and pilot plant runs
- Comparison of PDU and pilot plant data to determine the variation, if any, in H-Coal process performance at these two scales of operation
- Incorporation of the process performance variations in the Phase 0 commercial design
- Assessment of the economic impact, if any, resulting from incorporating the process performance-related changes in the commercial design

Since the work performed in this study is closely related to MITRE's recently completed H-Coal environmental control technology assessment study⁽²⁾, we relied on this previous work to obtain information on H-Coal process description and status. Additional information and data sources are listed in Table 1.

TABLE 1

INFORMATION/DATA SOURCES FOR THE STUDY

- BACKGROUND AND CONCEPTUAL DESIGN INFORMATION
 - Published reports on DOE-, EPA-, and EPRI-sponsored studies
 - Technical papers presented at recent conferences
- PROCESS OPERATING DATA
 - Reports published on PDU operation
 - Monthly progress reports of Catlettsburg pilot plant
 - Telephone contacts with engineering personnel at the pilot plant
 - Site visits to the pilot plant
- DESIGN AND COST DATA FOR THE COMMERCIAL PLANT
 - Blue books developed for Phase 0 design of Breckinridge commercial plant
 - Detailed process flow diagrams of Breckinridge project
 - Telephone contacts with ASFI, Bechtel Petroleum, Inc., and HRI project engineers for Breckinridge project

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2.0 H-COAL PROCESS DESCRIPTION AND DEVELOPMENT PATH

2.1 Process Description

The H-Coal process is a direct catalytic hydroliquefaction process based on the use of an ebullated-bed reactor for the conversion of coal to liquid products.⁽²⁾ Depending on process parameters, the H-Coal process can produce a broad spectrum of hydrocarbon liquids ranging from a low-sulfur fuel oil to an all-distillate syn-crude. The ebullated-bed H-Coal technology was a modification of the H-Oil reactor developed by Hydrocarbon Research, Incorporated (HRI).

The heart of the process is the ebullated-bed reactor shown schematically in Figure 1.⁽³⁾ The reactor feed, a coal/process-derived oil slurry, and a recycle stream from the ebullating internal pump, enter the bottom of the reactor through a distribution plate. The coal slurry has already been partially dissolved in the preheater. The liquid upflow velocity causes the bed of catalyst to fluidize. The catalyst is usually 1/16-inch diameter aluminum-oxide-based extrudates impregnated with the sulfides of cobalt or nickel and molybdenum. The catalyst level remains in the bed below the internal suction cup, while the reactor products, including the unconverted coal and mineral matter, leave the bed and are processed in a series of let-down vessels that separate gases, light products and heavy products, including unreacted solids. Because the catalyst is in a state of constant fluidization, portions of the catalyst can be continuously withdrawn and fresh catalyst can be

FIGURE 1
H-COAL EBULLATED BED REACTOR

added without interruption of the process to maintain an overall uniform level of catalytic activity. The ebullated bed allows intimate contact of catalyst, hydrogen, and coal solution, and achieves essentially isothermal reaction conditions and a constant reaction differential pressure. The continuous liquid phase in the reactor assures a good heat sink to provide reactor stability.

A schematic of the H-Coal process concept for the syncrude mode of operation is shown in Figure 2.⁽³⁾ The coal/process oil slurry (about 33-percent coal) is preheated to about 750-800°F under hydrogen pressure and is introduced into the H-Coal reactor with more hydrogen. Total reactor effluent vapor is separated from effluent slurry in the reactor effluent separation system, which operates at reactor conditions. The reactor slurry is let down and sent to hydroclones for partial solids separation. Hydroclone overflow is recycled back to the slurry preparation unit for supplying the major part of the recycle oil component. The underflow is vacuum-flashed to produce vacuum bottoms containing primarily mineral matter, unreacted coal (IOM), a small amount of 975°F distillate, and the overhead is combined with the other liquid distillate and fractionated. A portion of the fractionator bottoms is used as recycle solvent, and the remainder is distillate oil products. The overhead vapors from the H-Coal reactor are separated in a gas-liquid separator, the distillate is fractionated, and the gases are cleaned. The purified hydrogen is recycled to the front end and the hydrocarbon gases are products. During balanced

Source: Reference 3

FIGURE 2
SCHEMATIC OF THE H-COAL PROCESS
FOR THE SYNCRUDE MODE

operation, the vacuum bottoms produced are sufficient to provide the net hydrogen requirement for the liquefaction reaction.

2.2 History of H-Coal Process Development

The H-Coal process development efforts are presented in Table 2. Development of the process began in 1965, progressing through bench-scale testing, process development stages, and successful pilot-scale operations, to the preliminary design of a first commercial plant. The bench-scale coal liquefaction unit handle 25 pounds of coal per day. These bench-scale units are utilized for process improvement studies, including catalyst evaluation and for testing new coals. The 3.5-TPD capacity PDU has been used primarily for longer test runs and to simulate large-scale operations to confirm the design basis and operating parameters for the large Catlettsburg pilot plant.

Over twenty different coals have been tested in more than 60,000 hours of combined operation of the bench-scale and PDU equipment. These have included coals from the eastern United States, Texas and North Dakota lignites, western United States coals, Australian brown coal, German Steinkohle, South African coals, and coals from the United Kingdom. A successful longer term, about 30 days, PDU operation (PDU Run 5) on Illinois No. 6 coal has established the design and operating parameters for the 200-TPD capacity pilot plant in Catlettsburg, Kentucky.⁽⁴⁾ The PDU Run 5 was an important milestone in the H-Coal process development efforts. During this run, the operation of catalyst addition-

TABLE 2
HISTORY OF H-COAL PROCESS DEVELOPMENT

Period of Active Operation	1963 to 1968*	1969 to 1978	1980 to 1982	Proposed Startup 1989
Stage of Research and Development				
	Bench-Scale Research			
		Process Development Unit Operations		
		Lignite Wyodak Illinois No. 6 Kentucky No. 9, 11	Pilot Plant Operations	
			Illinois No. 6 Kentucky No. 9, 11 Wyodak	Preliminary Design for Commercial Plant
Coal Types:	Twenty Different Coals, Ranging from Lignite to Bituminous			
Capacity:	25 lbs/day	3.5 tons/day	200-600 tons/day	16,500 tons/day
C ₄ *-distillate** Output:	---	10 Barrels/day	628 Barrels/day	45,825 Bbls/day
Hours of Operation	50,000	14,000	7,000	Proposed Commercial Startup, 1988

*This unit is still used for testing different coals and process variations.

**Typical for Illinois coal.

withdrawal system, a key feature of H-Coal ebullated-bed technology, was successfully demonstrated for the first time. Also, the product yield structure obtained during this run provided the basis for the preliminary design of a proposed commercial plant.

The major objectives of the Catlettsburg H-Coal pilot plant were to provide data for the final design, construction, and operation of a commercial plant. The pilot plant was put into operation in March, 1980, with operation on coal commencing in May, 1980. During the two years of operation between May, 1980 and August, 1982, the pilot plant has logged approximately 7,000 hours of successful operation, while processing three coals (Illinois No. 6, Kentucky 9/11, and Wyodak. The longest pilot plant run (Run 8) lasted 131 days, during which time an average of 214 tons per day of Illinois No. 6 coal were processed. During this run, 19,200 tons of coal were converted to low-sulfur distillate at a yield of three barrels of oil per ton of dried coal.⁽⁵⁾ Run 8 has demonstrated the mechanical operability and reliability of critical equipment. In addition, the material balance data collected during Run 8 has essentially confirmed the product yield structure obtained during PDU Run 5. However, there were some differences in hydrogen consumption and yield of propane, butane, naphtha, and residuum fractions. These differences are discussed in detail in a later section of this report.

The first commercialization efforts for the H-Coal process began in late 1978, when Ashland Synthetic Fuels, Inc. (ASFI) submitted an unsolicited proposal to the U.S. Department of Energy for a Phase 0 cooperative effort for a detailed process design of a commercial-scale H-Coal liquefaction plant. In December, 1979, the decision was made to site the plant at Breckinridge, Kentucky. The proposed commercial plant was to convert about 23,000 TPD of run-of-mine coal into approximately 50,000 BPD of hydrocarbon liquids. The original Phase 0 design was to process a mixture of feedstocks that included Illinois No. 6, Kentucky No. 9, and Kentucky No. 11 coals. As indicated earlier, the operating results of PDU Run 5 were used as the basis for the Phase 0 design.

The first attempt to obtain support for the Breckinridge project from the Synthetic Fuels Corporation was unsuccessful because of insufficient equity participation. Cost escalations in the capital requirement for the Phase 0 plant resulted in a decision to scale down the plant size to approximately one-half of the original design. The SFC solicitation for the revised, or scaled-down, Phase 0 design was dropped when Ashland pulled out of the project. Bechtel, Inc. has resubmitted the Breckinridge project in the third SFC solicitation. However, the project was again dropped because of the lack of requisite equity participation.⁽⁶⁾

The pilot plant at Catlettsburg has recently been sold to ASFI, with the understanding that ASFI keep the plant in operational condition for the next three years.

3.0 DESCRIPTION OF H-COAL PDU AND PILOT PLANT FACILITIES AND OPERATIONS

The data collected from PDU operations in conjunction with additional process and equipment performance-related information generated during the pilot plant operations provide a solid basis for the design, construction, and operation of a commercial-scale plant. In order to facilitate the evaluation of operational data, brief descriptions of H-Coal process development and pilot facilities are given below.

3.1 H-Coal Process Development Unit

A schematic flow diagram of the H-Coal process development unit is given in Figure 3.⁽⁷⁾ The PDU, as indicated earlier, is used for intermediate-scale H-Coal process development efforts. It has a nominal capacity of approximately three tons of coal per day. The heart of the PDU unit is an ebullated-bed reactor which is used for the catalytic hydroconversion of slurried coal to distillate quality liquid products. The PDU contains all the required unit operations, i.e., coal feed preparation and slurry preparation, reactor product separation, etc., which are duplicated in the larger capacity pilot plant at Catlettsburg, Kentucky. The PDU can be divided into the following functional areas.

- Coal Handling Section
- Feed Slurry Preparation Section
- Reactor Section
- Product Separation Section

Source: Reference 7.

FIGURE 3
SCHEMATIC FLOW DIAGRAM OF H-COAL
PROCESS DEVELOPMENT UNIT

3.1.1 Coal Handling Section

The coal handling section, not shown in Figure 3, receives coal in rail cars (50- to 80-ton capacity) as 1-1/2" lumps. The coal is unloaded and transferred to a crushing area by using belt conveyors. In the crushing area, its size is reduced to less than 1/2", and the crushed coal is then transferred via a bucket elevator into the storage hoppers located above the roof.⁽⁷⁾ The coal pulverizers, situated directly under the coal hoppers, are gravity-fed, and convert the crushed coal from less than 1/2" particle size to 95 percent through 60-mesh. The powdered coal is then transferred to a storage hopper which is located directly above a rotary tray drier.

In the tray dryer, the pulverized coal is dried under a nitrogen atmosphere at about 210-220°F. The normal coal residence time in the dryer is about 35 minutes.⁽⁷⁾ The dried (containing about 2 percent moisture), pulverized coal is transferred to a shaker screen via a water-cooled screw conveyor. The properly sized, dry coal powder is transferred under a nitrogen blanket to an intermediate storage tank. The feed coal is now ready to be transferred, via a feed metering hopper, to the slurry mix tank.

3.1.2 Feed Slurry Preparation Section

The pulverized coal is continually fed from the metering hopper into a slurry mix tank where it is mixed with clean recycle oil stream from the slurry oil holding tank and hydroclone overflow (see Figure 3). The clean recycle oil stream primarily consists of

atmospheric still bottoms. However, during periods of solvent imbalance, vacuum still overhead and external solvent are used. The coal-oil slurry from the mixing tank is transferred to a feed preheater by using high-pressure feed pumps. The mixing action in the slurry tank is provided by a mixer and the recirculating pumps, which also maintain positive, high-velocity suction to the high pressure feed pumps. A separate high-pressure oil pump is kept as a standby unit to come on-line in case of feed pump failure or reactor emergency.

3.1.3 Reactor Section

The coal-oil slurry is pumped to reactor pressure and mixed with makeup hydrogen stream before entering the gas-fired feed preheater. After preheating, the slurry-hydrogen mixture is fed into the bottom head of the H-Coal reactor. A recycle gas stream is taken from the top of the reactor, and it is scrubbed with oil to remove higher hydrocarbons. The scrubbed gas stream is recompressed and passed through a smaller preheater, then introduced into the bottom of the H-Coal reactor below the coal-slurry feed point. The recycle gas and slurry feed temperatures are controlled manually.

A Chempump is used to maintain the recycle oil flow within the reactor which in turn controls the catalyst-bed expansion. The oil flow is manually controlled by adjusting the frequency of pump motor and control valve setting to maintain a desired catalyst bed level within the reactor.⁽⁷⁾ Suction for the Chempump, or ebullating pump, is taken through an internal recycle cup which is located

above the catalyst bed level (see Figure 1, Section 2.0). Flush oil is used to keep the rotor and bearing areas of the Chempump unit free of solids.

The reaction temperature is monitored by a ten-point, internal thermocouple assembly. The reactor wall is wound with resistance wire and insulated.

3.1.4 Product Separation Section

The reactor liquid-vapor effluent passes through an external separator where the vapors are separated from the liquid effluent and are sent to a scrubbing system. The liquid or slurry portion of the reactor effluent is sent to either one of two parallel pressure letdown valves where the effluent slurry pressure is brought down to 100 psig. The slurry is then sent to an atmospheric flash vessel where it separates into vapor and liquid-solid streams.

The vapors from the atmospheric flash drum are passed through a condenser and then into a vapor/liquid separator. The noncondensable portion of the flash vapors is vented overhead of the separator, and after metering, sent to a flare system. The condensed liquid from the separating vessel is pumped through a preheater into the atmospheric still.

The solids-containing liquid stream taken from the bottom of the atmospheric flash vessel passes through a control valve and enters a hydroclone system.* The hydroclone overflow is recycled to

*Depending on the mode of operation, the solids-containing stream can be sent directly to a vacuum still.

the slurry mixing tank, while the solids-enriched underflow is fed to a vacuum still. The temperatures in the feed lines and vacuum still are maintained by electrical resistance heat and are constantly monitored by internal and external thermocouples. Steam ejectors are used to create a vacuum inside the tower which operates at an absolute pressure of 14 mm. mercury. The vacuum still overhead is condensed and collected in an overhead receiver. The overhead product is transferred by centrifugal pump to a weigh station and then sent to the clean oil storage tank in the slurry preparation section. The vacuum still bottoms are collected in open-top drums blanketed with CO₂.

The reactor vapors pass through a water-cooled heat exchanger, and, after being cooled to ambient temperature, enter the overhead separator. Water is continuously injected before the heat exchanger and after the overhead separator to dissolve the water soluble salts which can plug the heat exchanger and other transfer lines. The uncondensed gases from the overhead separator are sent to a scrubber where they are scrubbed by a counter-current stream of No. 2 fuel oil. This scrubbing will remove hydrocarbon gases and some hydrogen sulfide from the hydrogen-rich recycle gas. The absorbed gases are released from the scrubber oil in the flash drum and are flared. The lean oil is pumped back to the scrubber. The hydrogen-rich recycle gas from the top of the scrubber passes to a knockout drum before being recompressed to the reactor pressure. The purity of

the recycle gas is controlled by the circulation rate of scrubber oil and the quantity of recycle gas vented from the knockout drum.

The condensed liquid from the reactor overhead separator is let down to a pressure of about 100 psig by using one of the two letdown valves and sent to a water separator vessel. The hydrocarbon liquid and flash vapors are sent to the atmospheric still. The atmospheric still overhead product, along with flash gas and water vapor, pass to the overhead receiver. The net overhead product is separated and collected in a separate product storage drum. The sour water passes to an outside storage tank via another intermediate storage drum. The flash gas is metered and flared. The bottoms from the still are water-cooled and sent into an intermediate storage drum. A part of the atmospheric still bottoms is sent to the "clean slurry oil" holding tank in the slurry preparation section. The rest of the bottoms are collected as products.

The atmospheric still is a ceramic ring-packed, electrically heated tower.

3.2 H-Coal Pilot Plant Facility

The liquefaction and oil recovery section of the H-Coal pilot plant, while functionally similar to the PDU unit described earlier, is more elaborate, larger in size, and closely duplicates the expected operation of the single train of the H-Coal commercial plant. The H-Coal pilot plant was designed for two modes of operation: (1) a low-throughput, high-severity operation yielding all-distillate synthetic crude oil as the main product (Syncrude

mode), and (2) a high-throughput, low-severity operation yielding low-sulfur fuel oil as the main product (fuel oil mode).

The capacity of the plant in the first mode of operation, i.e., high-severity, low-throughput, is about 220 short tons of as-received coal per day. In the fuel oil mode, the plant capacity increases to approximately 630 short tons of as-received coal per day. The pilot plant was only operated in the syn crude mode, because of fund limitations.

The pilot plant incorporates the equipment necessary to demonstrate the H-Coal concept with respect to slurry preparation, preheating, and liquefaction. Product recovery and separation facilities allow the reactor effluent to be processed and separated into gaseous, liquid, and slurry streams. The hydrogen-rich gas effluent is cleaned and recycled for reuse. The hydroclone overflow stream is recycled as the major portion of the recycle oil, the underflow is vacuum-flashed to produce oil, and bottoms are flaked and disposed of as a shiny black solid.

The pilot plant is located adjacent to the Ashland refinery and advantage is taken of the existing refinery facilities to provide support to the H-Coal facility. This support includes the supply of hydrogen, a part of process steam and utilities, and processing of sour gases. The pilot plant has its own wastewater treatment facility, which treats liquid effluents before discharge to the river. In addition, the pilot plant has its own utilities and waste oil recovery sections.

A schematic of the liquefaction and product oil recovery section of the pilot plant is given in Figure 4.⁽⁸⁾ At the pilot plant, washed coal up to 4 inches in size is received by rail cars and/or trucks.* The unloaded coal is stored in stock piles, from where it is reclaimed with front-end loaders, and crushed to -3/4-inch size before storing in holding bins. The coal feed is further ground in a bowl mill to a -100-mesh size, and dried to 2-percent moisture, and then moved to storage bins.⁽⁹⁾

Dried, pulverized coal is mixed with recycle hydroclone overflow and fractionator bottoms in the slurry mix tank (Q236).** The slurry transfer pump provides the feed coal-oil slurry to the suction side of the high-pressure slurry feed pumps (J219).** The slurry charge or feed pumps bring the pressure of the slurry slightly over 3,000 psig. A portion of recycle hydrogen is mixed with the slurry and passed through a feed slurry preheater, where its temperature is increased to about 720°F. A separate hydrogen preheater is provided to heat the rest of the recycle hydrogen stream to a temperature of about 750°F. The hot slurry and hydrogen streams are fed to the bottom of the H-Coal ebullating-bed reactor (K201).

*The Coal receiving and preparation plant is not shown in Figure 4.

**Corresponds to equipment numbers shown in Figure 4.

FIGURE 4
SIMPLIFIED H-COAL PILOT PLANT SCHEMATIC DIAGRAM

The effluent leaving the reactor is first passed to the high-pressure (about 3,000 psig) reactor effluent separator (Q224), where the gas is separated from the slurry. The gas leaving the separator is cooled in a trim cooler to about 550°F.⁽⁹⁾ The partially cooled stream is passed to a vapor/liquid separator, where the condensed liquid is separated from the gas. The gas from the separator is further cooled to 130°F in an overhead cooler. Before the gas enters the overhead cooler, a direct water injection is used to dissolve the soluble salts which might otherwise plug the heat exchangers and other transfer lines. The cooled effluent vapor/liquid stream is sent to a separator where condensed hydrocarbon liquids, water, and gas are separated. The hydrogen-rich, noncondensable gas fraction is scrubbed in a lean oil absorber system at 3,000 psig, to remove light hydrocarbon gases. The scrubbed hydrogen-rich gas mixture (absorber overhead stream) is recompressed and sent to the hydrogen preheater. The oil from the absorber is flashed to 50 psig to release the hydrocarbon gases. The lean oil is then pumped to 3,000 psig and recycled to the absorber.

The liquid (or slurry) effluent stream from the high-pressure separator goes through two stages of pressure letdown to reduce the pressure from 3,000 psig to approximately 70 psig. In the first stage pressure letdown, the slurry pressure is brought down to 1,200 psig, and the throttled slurry is sent to the intermediate pressure separator, where the gases are separated from the slurry. The gases from the intermediate separator are partially cooled and sent to a

separator. The vapor stream from this separator is washed with direct water injection to remove dissolved salts, and passed through a final cooler, which brings the gas temperature to approximately 125°F. The cooled vapor/gas stream flows to a separator, where the hydrocarbon liquid, water, and hydrogen-rich gas are separated. The gas is then combined with the other make-up hydrogen stream.

The slurry from the intermediate stage letdown separator vessel is throttled to a pressure of approximately 70 psig, and sent to a flash drum. The gases produced in the low-pressure flash drum are also cooled, in two steps, to a temperature of approximately 125°F, with intermediate product separation and direct water injection, as described above. The gas separated from the low-pressure system is vented to the plant fuel gas system. All distillates condensed and separated at 3,000, 1,200, and 70 psig pressures are transferred to the fractionator feed drum.

The slurry from the low-pressure separator is sent to a hydroclone system where a partial separation of unconverted coal and ash is carried out. The hydroclone overflow, after atmospheric and vacuum flashes, is used to provide a major portion (about 60 percent) of the recycle oil in the slurry preparation section. The lower boiling fractions separated from the hydroclone overflow are sent to the fractionating tower. The underflow from the hydroclones, partially enriched in coal and ash, is heated and sent to an atmospheric stripper, followed by a vacuum stripper, to remove distillate oil from the slurry. (9)

The bottoms from the vacuum stripper contain all the residual (975+°F) material, unconverted coal and ash. This material is cooled and solidified (or flaked) on stainless steel belts to facilitate transportation and disposal. During the PDU operations, this material was collected in open top drums, under a CO₂ blanket to minimize oxidation. In the commercial H-Coal plant, the vacuum bottoms will be gasified to produce hydrogen.

All the distillate streams recovered in different sections of the plant are sent to the main fractionator N-202. The overhead from this fractionator is an unstabilized naphtha stream which is sent to a stabilizer, and then to product storage. A side draw from the main fractionator is stripped to remove light hydrocarbons; the light distillate product from the bottom of this side stripper is cooled and sent to the product storage area. A small fraction of this light oil is returned to the main fractionator. A portion of the main fractionator bottoms, i.e., heavy oil, is used to supply part of the recycle solvent in the slurry preparation section. The balance of this heavy oil is sent to the product storage area.

3.3 Description of Selected PDU and Pilot Plant Runs

In order to assess the performance of the H-Coal process at two scales of operation, specific PDU and pilot plant runs were selected. The selected runs achieved smooth plant operations with sustained steady-state product yields. Projected H-Coal performance was more or less duplicated during these runs. PDU Run 5 and Pilot Plant Run 8 are described below.

3.3.1 PDU Run 5

The PDU Run 5 was made in the syncrude mode, high severity, low throughput operation, with Illinois No. 6 coal as the feedstock. During this run, the PDU operated continuously for 717 hours (30 days) from September 4, 1977 to October 5, 1977, with successful on-line catalyst addition and withdrawal. The catalyst addition and the withdrawal system were successfully demonstrated for the first time in the PDU test program. An average replacement rate of 1.07 pounds of catalyst per ton of coal feed was maintained during the run.⁽⁴⁾ A coal throughput of 31 pounds per hour per cubic foot of reactor, which is a typical value for the syncrude operation, was consistently achieved during the run. The PDU reactor volume in the syncrude mode is about 7.5 cubic feet.⁽⁴⁾ Therefore, the average coal feed rate to the reactor was maintained at about 232 pounds of coal per hour.

During the first ten days of the run, the distillate yields were lower than the residual oil yields.⁽⁴⁾ The lower distillate oil yields are attributed to lower hydrogen partial pressure and decreased concentration of residual oils in the recycle slurry. As the hydrogen partial pressure and residual oil content of the recycle slurry were increased after the 15th day of PDU operation, the distillate oil yield increased, with a corresponding decrease in the residual oil yield.

A quasi-steady-state with respect to catalyst activity was reached in about ten days. Equilibrium catalyst activity is

achieved when the addition and withdrawal of catalyst maintains a constant residuum yield.⁽⁴⁾ This equilibrium activity was obtained during the final days (the 29th day onward) of the PDU operation. The yield of residuum obtained during this period closely approached the equilibrium yield predicted for ideal operating conditions, i.e., constant feed rate, recycle rates, temperature, and hydrogen partial pressure.⁽⁴⁾

PDU Run 5 also successfully demonstrated the use of petroleum-based oil as startup oil, instead of anthracene oil used in the earlier PDU runs. Also, the results of PDU Run 5 were in better agreement with the bench-scale runs and the pilot plant design basis when compared to the earlier PDU runs (Runs 1 through 4). The PDU was successfully able to simulate the actual operating conditions expected at the Catlettsburg pilot plant. The overall product yields obtained during the 29th day (or period 29) of the PDU runs were used to develop Phase 0 design of the proposed H-Coal commercial plant. The Phase 0 design basis will be discussed in a later section of this report.

3.3.2 Pilot Plant Run 8

During Run 8, the H-Coal pilot plant in Catlettsburg, Kentucky operated continuously for 131 days in the syncrude mode. The run was started on August 7, 1981 and was concluded on December 11, 1981.⁽⁵⁾ The run was initiated with Kentucky No. 9 coal, but later switched to Illinois No. 6 coal. During the 131 days of

continuous operation, the plant was operated on coal for 89.7 days, thus giving an on-stream factor with coal of 71.6 percent. When the coal feed to the reactor was interrupted, the backup oil pumps would come on-stream to maintain the plant operation.

The plant target coal feed rate was set at 219 tons of dry coal per day. The average feed rate over the coal operating period was about 214 tons of dry coal per day. Within the first few days of operation, a space velocity of 31 pounds of coal/hr/ft³ of the reactor volume was established. This space velocity was maintained throughout the run, except when equipment operational problems warranted reduction in the coal feed rate. A total of 22,000 tons of as-received coal, or 19,200 tons of dry coal, was processed.

A total of 16,885 pounds of catalyst was used during Run 8. This is equivalent to replacing the initial catalyst inventory approximately 2.3 times.⁽⁵⁾ The average catalyst replacement rate was approximately one pound of catalyst for every ton of coal fed to the H-Coal reactor.

A special four-day material balance run was conducted to improve the accuracy of overall product yield data from the pilot plant. This material balance run was made during the period from October 29 through November 2, 1981. During this period, the plant was operating closer to the operating conditions of the proposed commercial plant. Efforts were made to account for all possible product losses, and to maintain the plant operations at steady-state conditions. A material balance closure of 96 to 98 percent was obtained.

Run 8 has demonstrated a successful scale-up of the ebullated-bed reactor system from 3 TPD to 220 TPD of dry coal capacity. In addition, the following objectives were achieved:⁽⁵⁾

- Demonstrated operations in solvent balance
- Demonstrated good hydroclone efficiency
- Demonstrated the use of the lean oil system for recycle hydrogen purification
- Achieved an overall product yield comparable to PDU Run 5

A detailed comparison of the H-Coal process performance at the two scales of operation is given below.

3.4 Comparison of H-Coal Performance at PDU and Pilot Plant Scale Operations

Specific periods of operation during PDU Run 5 and Pilot Plant Run 8 are selected to illustrate the H-Coal process performance. For PDU Run 5, Period 29 was selected to represent conditions close to equilibrium with respect to catalyst activity and product yields. The special four-day material balance period of Pilot Plant Run 8 was selected to represent comparable equilibrium conditions. Comparison of the operating conditions and results follows.

3.4.1 Comparison of Operating Conditions

Both PDU Run 5 and a large part of Pilot Plant Run 8 were made with Illinois No. 6 coal from the Burning Star Mine. The average characteristics of coal feedstock used during these runs are presented in Table 3.^(4, 5) The average carbon content of coal feed used in PDU Run 5 is about 5 percent higher than that used in Pilot Run 8. The nitrogen content of the coal used in PDU Run 5 is about

TABLE 3

CHARACTERISTICS OF FEEDSTOCK USED IN
SELECTED PDU AND PILOT PLANT RUNS

	<u>PDU - Run 5</u>	<u>Pilot Plant - Run 8</u>
Coal Type	Illinois No. 6 - Burning Star Mine	
Moisture Content, Wt. % (Feed Coal)	3-4	2
<u>Proximate</u> , Dry Basis, Wt. %		
Ash	10.61	11.80
Volatile Matter	37.90	38.8
Fixed Carbon	51.49	49.4
<u>Ultimate</u> , Dry Basis, Wt. %		
Carbon	70.88	67.45
Hydrogen	4.91	4.93
Nitrogen	1.47	1.28
Sulfur	3.47	3.6
Ash	10.61	11.75
Chlorine	0.07	0.1
Oxygen	8.57	10.89

14 percent higher than that used in Pilot Run 8. The chlorine and oxygen contents of coal used in Pilot Run 8 are higher than that used in PDU Run 5 by approximately 43 percent and 27 percent, respectively.

All the elemental composition comparisons are done on dry basis. A particle-size distribution for the coal feed used in the selected PDU and pilot plant runs is given in Table 4. Approximately 88 percent of the coal feed was -100 mesh size. The PDU feed contained approximately 64 percent more -325 mesh material than the pilot plant feed.

The properties of fresh catalyst used in PDU Run 5 and Pilot Plant Run 8 are compared in Table 5. The physical properties of the catalyst used in these two runs are very similar. The chemical properties of the catalyst, however, are slightly different; the pilot plant catalyst batch contained higher moly and cobalt content than the one used in the PDU run.

The operating conditions of the PDU and the pilot plant during the selected respective periods of operation are presented in Table 6.^(4, 5) The operating conditions during these specific periods of operation are quite comparable, except for the reactor operating pressure and cumulative catalyst age. The reactor operating pressure during the pilot run was about 3000 psig, whereas operating pressure of the PDU was about 2700 psig. A cumulative catalyst age of 1695 lb of coal/lb of catalyst was obtained for the pilot run as opposed to 1,102 lbs coal/lb catalyst for the PDU run,

TABLE 4

PARTICLE SIZE OF COAL FEED USED IN
SELECTED PDU AND PILOT PLANT RUNS

<u>Screen Analysis</u>	<u>PDU - Run 5</u>	<u>Pilot Plant - Run 8</u>
	<u>Weight Percent</u>	
+50	2.59	0.37
50/70	2.75	2.59
70/100	5.82	8.93
100/140	8.68	18.41
140/200	10.31	18.19
200/325	23.42	23.23
325/Pan	<u>46.43</u>	<u>28.28</u>
TOTAL	100.00	100.00

TABLE 5

PROPERTIES OF FRESH CATALYST USED IN SELECTED
PDU AND PILOT PLANT RUNS

	<u>PDU - Run 5</u>	<u>Pilot Plant - Run 8</u>
TYPE	American Cyanamid HDS 1442A	
Bulk Density lb/ft ³	37	37
Crush Strength lb/m.m	3.1	3.6
<u>Chemical Composition (Wt. % Dry)</u>		
Molybdenum	9.35	10.33
Cobalt	2.4	3.8
Titanium	---	---
Boron	---	---
Iron	---	0.03
Al ₂ O ₃	---	---
Na ₂ O	0.02	0.02
Extrudate Diameter, mm.	1.6	1.57
Extrudate Length, avg. mm.	4.3	5.0
Pore Volume (m ³ /gm)	0.72	0.74
Surface Area (m ² /gm)	355	350

*Compacted.

TABLE 6

COMPARISON OF OPERATING CONDITIONS FOR SELECTED
PDU AND PILOT PLANT RUNS

Run Designation Parameter	PDU Run 5, Period 29 October 2-3, 1977	Pilot Plant Run 8, 4-Day Mat'l. Bal. Period Oct. 29-Nov. 2, 1981
Average Reactor Temp., F°	851	848
Reactor Pressure, PSIG	2,700	3,006
Hydrogen Partial Pressure, PSIG		
• Inlet	2,490	2,512
• Outlet	1,920	2,058
Cumulative Catalyst Replace- ment Rate, lb/ton dry coal	1.06	0.96
Cumulative Catalyst Age, lb coal/lb catalyst	1,102	1,695
Coal Space Velocity, lb/hr/ft ³	31.46	29.2
Recycle Slurry Composition		
• Hydroclone Overflow	65 %	58 %
• Atmospheric Still bottoms	35 %	42 %

thus indicating that the pilot run was closer to the catalyst equilibrium activity. The recycle slurry contained from 58 to 65 percent (on solid free basis) of hydroclone overflow, the balance being the atmospheric still bottoms.

3.4.2 H-Coal Process Performance Comparison

The products yield and quality are used as the major indicators of H-Coal process performance at the two scales of operation. The overall product yield obtained during the selected PDU and pilot plant run periods is given in Table 7.^(4, 5) The overall distillate oil (C_4 -975°F) yield obtained during the PDU and the pilot runs are in excellent agreement. The C_4 -975°F distillate oil yields (on a weight basis) for both runs were higher than 47 percent of the dry coal fed to the H-Coal reactor. However, the pilot plant produced approximately 81 percent more C_4 fraction on a weight basis than the PDU. The yield of C_4 -distillate oil on a volumetric basis is approximately 5 percent higher for the pilot plant run than the PDU run. The C_3 fraction yield obtained in the pilot run is also significantly higher, about 34 percent, than the PDU run. Because of the higher C_3 and C_4 fraction yields the overall hydrogen consumption has increased from 4.9 percent of the dry coal input for PDU run to 5.47 percent of the dry coal for pilot run, an increase of about 11 percent.

The yield of naphtha fraction, C_5 -400°F, is approximately 10 percent higher for the pilot run when compared to the PDU yield. The yield of 400-650°F oil fraction is roughly 20-percent lower

TABLE 7

COMPARISON OF OPERATING RESULTS FROM
SELECTED H-COAL PDU AND PILOT RUNS

PLANT OUTPUT	Run Designation				
	PDU		PILOT PLANT		
	Run 5, Period 29		Run 8, 4-day Material Bal.		Run 11*
	LBS/100 LBS OF DRY COAL	BBL/TON OF DRY COAL	LBS/100 LBS OF DRY COAL	BBL/TON OF DRY COAL	LBS/100 LBS OF DRY COAL
H ₂	-4.91		-5.47		-4.74
H ₂ S	2.64		2.67		2.63
NH ₃	1.08		0.65		1.08
H ₂ O	6.67		6.67		6.67
CO ₂	0.48		—		0.48
C ₁	3.75		4.20		3.75
C ₂	3.54		3.02		2.94
C ₃	3.39		4.55		3.20
C ₄	2.54		4.61		1.88
C ₅ - 400°F	16.20	1.40	17.8	1.69	16.67
400- 650°F	20.37	1.21	16.46	0.99	19.30
650- 975°F	7.96	0.41	8.81	0.46	10.39
Residuum	19.00		21.26		21.68
Unconverted Coal	5.78		3.46		3.51
Ash	11.51		11.31		10.58
TOTAL:	100.00		100.00		100.00
Total Distillate, C ₄ ⁺ - 975°F	47.1	3.01	47.7	3.14	48.24
Residuum, Unconverted Coal and Ash	36.3		36.0		35.8

*Mr. Richard Santore of PETC, in reviewing this manuscript, called attention to the results of Pilot Plant Run No. 11. The results of this run agree more closely with PDU Run No. 5 in terms of hydrogen consumption. The conclusions of this report with respect to the negligible impact of deviations in Run No. 8 from PDU No. 5, are still unaltered.

for the pilot run when compared to the yield obtained during the PDU run. The yield of 650-975°F fraction is roughly 10 percent higher for the pilot run than the corresponding PDU run. Overall, during the pilot plant run, higher yields of lighter products, i.e., C₃, C₄, and naphtha, and slightly lower yields of 400-650°F fractions were obtained when compared to the PDU yields. Finally, the residuum plus unconverted coal and ash fractions are in excellent agreement between the PDU and pilot plant runs.

The elemental composition and API gravity of naphtha (180-380°F) and light distillate oil (380-650°F) product fractions are compared in Table 8.⁽⁵⁾ The naphtha fraction obtained in the pilot run is slightly lighter than that obtained in the PDU run; it contains slightly higher quantities of carbon and hydrogen. The nitrogen content of the naphtha fraction produced in Pilot Run 8 is significantly lower, about 61 percent of the PDU value, than that produced during PDU Run 5. The higher nitrogen content in the PDU samples may be due partly to the high nitrogen content of the coal used during the PDU run (See Table 3). The oxygen content is significantly higher in the sample collected from the pilot run than the PDU sample. The oxygen discrepancy may be due to the different analytical techniques used for its measurement. Also, the difference in oxygen content of the coal may be partly responsible for the oxygen discrepancy.

As was the case with the naphtha fraction, the light distillate oil fraction obtained from the pilot run is slightly lighter than

TABLE 8

PROPERTIES OF NAPHTHA AND LIGHT DISTILLATE OIL
FOR SELECTED PDU AND PILOT RUNS

RUN DESIGNATION :	PDU	PILOT PLANT
	RUN 5, PERIOD 29	RUN 8, 4-DAY MATERIAL BALANCE
<u>Naphtha (180-380°F)</u>		
Gravity, °API	36.5	39.6
Elemental Analysis (Wt. %)		
Carbon	85.14	86.78
Hydrogen	12.00	12.66
Nitrogen	0.31	0.19
Sulfur	-	0.26
Oxygen	0.83*	1.61
<u>Light Distillate Oil (380-650°F)</u>		
Gravity, °API	15.4	18.0
Elemental Analysis (Wt. %)		
Carbon	87.54	87.47
Hydrogen	9.57	9.82
Nitrogen	0.52	0.56
Sulfur	-	0.09
Oxygen	1.03	2.84**

*Unterzaucher.

**By neutron activation.

the similar fraction produced in the PDU run. The carbon, hydrogen and nitrogen concentrations are quite comparable in the light distillate oil fractions obtained from both PDU and pilot runs. Once again, a higher oxygen content has been indicated for samples produced in the pilot run when compared to similar samples obtained from the PDU run.

The elemental composition of heavy oil and vacuum bottoms fractions are presented in Table 9.⁽⁵⁾ Because the PDU heavy oil fraction has higher end point, 975°F, as opposed to 950°F, than the similar fraction from the pilot plant, its API gravity is significantly lower, thus making the heavy oil fraction produced during the PDU run significantly denser than that produced in the pilot plant. The carbon and hydrogen contents of the heavy oil fractions are quite comparable. The heavy oil fraction produced in the pilot run has slightly higher nitrogen content than the similar fraction produced in PDU Run 5.

The vacuum bottoms produced in the pilot runs contain slightly higher carbon content than that produced during the PDU run. The nitrogen content of the pilot plant sample was significantly higher than the sample produced during the PDU run.

Other pertinent properties of naphtha fractions collected from the PDU and pilot runs are given in Table 10.⁽⁵⁾ The refractive index (a measure of aromaticity) of the fractions produced during the PDU run is slightly higher than that produced during the pilot run, indicating that the PDU naphtha is slightly more aromatic than

TABLE 9

API GRAVITY AND ELEMENTAL ANALYSIS OF
HEAVY DISTILLATE AND VACUUM BOTTOMS

RUN DESIGNATION	PEU	PILCT PLANT
	RUN 5, PERIOD 29	RUN 8, 4-DAY MATERIAL BALANCE
<u>Heavy Oil Fraction (650-950°F)*</u>		
°API	-5.9	0.3
Elemental Analysis (Wt. %)		
Carbon	89.7	89.06
Hydrogen	7.13	7.09
Nitrogen	1.1	1.25
Sulfur	-	0.26***
Oxygen	0.97**	1.94***
<u>Vacuum Bottoms (950°F†)</u>		
°API	-	-
Elemental Analysis (Wt. %)		
Carbon	59.18	63.15
Hydrogen	4.08	3.93
Nitrogen	0.99	1.38
Sulfur	2.32	2.38
Oxygen	1.37	1.06
Ash	27.94	28.10

*For Pdu-5, cut was 650-975°F.

**Unteraucher.

***By difference.

TABLE 10

PROPERTIES OF NAPHTHA FRACTION (180°-380°F)

	<u>PDU</u>	<u>PILOT PLANT</u>
	RUN 5	RUN 8
Refractive Index (20°C)	1.47	1.459
Heating Value, Btu/lb	18,278	19,104
Bromine No., gms./100 gms	-	29.8
Phenols, Wt. %	9.2	6.49
Aniline Point	-	71.5
Research Octane No.	-	79.6
Existant Gum, mg/100 ml	-	46.8
Potential Gum, mg/100 ml	-	228.8

the pilot plant naphtha. The higher heating value of the pilot plant fraction is about 5 percent higher than the PDU fraction. The phenols content of PDU fraction is significantly higher, about 41 percent, than the pilot plant fraction. These differences may reflect the sharpness of the cutpoints in fractionating the liquids. For example, the next higher boiling fraction (400°F) contains a higher concentration of phenols. The other properties of middle oil fractions collected from the PDU and pilot runs are given in Table 11.^(4, 5) Except for the pour point, other physical properties, i.e, refractive index, flash point, and smoke point, are in excellent agreement. The pour point for the PDU fraction is about 15°F lower than the pour point for the pilot plant fraction. Other properties, such as heating value, bromine number, and phenols concentration, are in good agreement.

The other properties of heavy oil fractions (650-950°F) are compared in Table 12.^(4, 5) Since the fraction obtained from the PDU run has a higher end point, 975°F as opposed to 950°F, than the pilot plant fraction, its pour point and viscosity are higher than the fraction obtained from the pilot plant.

The properties of spent catalyst and the quality of the recycle solvent also provide indirect measures of the performance of the H-Coal liquefaction process. The analyses of spent catalyst from the selected PDU and pilot runs are given in Table 13.^(4,5) The average catalyst age, pounds of coal/pound of catalyst, was about 34 percent higher during the pilot plant run when compared to the

TABLE 11
OTHER PROPERTIES OF MIDDLE OIL FRACTION

	<u>PDU Run 5</u>	<u>Pilot Plant Run 8</u>
Refractive Index (20°C)	1.529	1.530
Pour Point, °F	-35	-20
Flash Point, °F	205	202
Viscosity, Centistokes		
100°F	6.0*	2.8
210°F	-	1.0
Heating Value, Btu/lb	17,706	17,947
Phenols, Wt. %	14.0	15.9

*At 72°F.

TABLE 12
OTHER PROPERTIES OF HEAVY OIL FRACTION (650-950°F)

	<u>Pilot Plant Run 8</u>	<u>PDU Run 5*</u>
Pour Point, °F	70	75
Flash Point, °F	410	--
Viscosity, Centistokes		
100°F	5,458	--
210°F	42.8	65
Heating Value, Btu/lb	17,059	16,742

*650-975°F Fraction.

PDU run. The coal operating period during PDU Run 5 lasted for approximately 735 hours, as opposed to 1,475 hours for Pilot Run 8. Both for the PDU and pilot run, over 99 percent of the catalyst particles were larger than 20-mesh, indicating that the catalyst particles were resisting any attrition resulting from the ebullating action within the reactor.

The carbon content of the oil-free, spent catalyst, or carbon deposition, was slightly higher, about one percentage point higher, for PDU Run 5 as compared to Pilot Run 8. The titanium content of the spent catalyst is significantly higher for samples collected from the pilot plant than the PDU. The high concentration of titanium has been observed in many coal liquefaction studies.

The recycle solvent properties for the selected PDU and pilot runs are compared in Table 14.⁽¹⁰⁾ The recycle solvent was made up of approximately 60- to 65-percent (by weight) hydroclone overflow (HO), and 35- to 40-percent atmospheric still or fractionator bottoms (FB). The hydroclone overflow fraction contains distillate oil, soluble residual oil, IOM, and ash, whereas the fractionator bottoms fraction is a totally distillate oil. The true boiling distillations for the composite fractions of the recycle solvent for PDU Run 5 and Pilot Run 8 are also presented in Table 14.⁽¹⁰⁾ The composite samples of the recycle solvent fractions are obtained by adding equal amounts of these fractions collected daily over specific periods of both PDU and Pilot Plant runs.⁽¹⁰⁾

TABLE 13

PROPERTIES OF SPENT CATALYST RECOVERY FROM
SELECTED PDU AND PILOT RUNS

	<u>PDU - Run 5</u>	<u>Pilot Plant - Run 8</u>
Average Catalyst Age (lbs coal/lb catalyst)	1,150	1,545
Time on Coal Operation	735	1,475
Benzene or Toluene (Soluble oil wt. %)	27.4	20.2
Pore Volume c.c/gm (Calculated)	0.416	---
Particle Size Distribution of Oil-free Catalyst:		
<u>Screen Analysis</u>		<u>Wt. %</u>
+20	99.4	99.6
20/30	0.12	0.05
30/50	0.+	0.01
50/100	0.14+	0.01
100/325	0.03	0.01
-325	---	.01
<u>Oil-free Catalyst Analysis</u>		<u>Wt. %</u>
Carbon	16.51	15.2
Molybdenum	6.15	8.1
Titanium	1.91	5.13
Iron	0.22	0.77
Calcium	0.14	0.01
Sodium	0.47	2.22
Cobalt	---	2.08

TABLE 14

COMPARISON OF SOLVENT PROPERTIES FOR SELECTED PDU AND PILOT PLANT RUNS

RUN DESIGNATION:	PDU Run 5, Period 29 October 2-3, 1977		Pilot Plant Run 8 4-day Material Balance Period Oct. 29 - Nov. 2, 1981	
	Percent	lb/lb solids	Percent	lb/lb solids
<u>Recycle Oil Composition</u>				
• Hydroclone Overflow (FO) Fraction	65	1.04	58	1.42
• Fractionator Bottoms (FB) Fraction	35	0.552	42	1.03
<u>Recycle Oil Fractions</u>	<u>Composite FB Fraction</u>	<u>Composite HO Fraction</u>	<u>Composite FB Fraction</u>	<u>Composite HO Fraction</u>
IBP °F	392	428	446	446
446°F Wt. %	0.20	0.1	0.0	0.0
446°F - 572°F	35.3	10.0	51.3	24.4
572°F - 662°F	32.7	26.0	29.2	31.7
662°F - 752°F	11.5	35.8	10.8	25.0
752°F ⁺	9.0	23.1	3.2	15.2
Cold Trap and Losses	11.30	5.0	5.5	3.7
TOTALS	100.	100.	100.	100.
<u>Elemental Composition</u>				
Carbon	90.64	91.	88.93	90.36
Hydrogen	8.79	7.88	9.13	8.44
Nitrogen	0.76	0.91	0.6	0.97
Sulfur	-	-	0.2	-

In both the PDU and pilot runs, the hydroclone overhead fraction is heavier than the corresponding fractionator bottoms fraction. The FB and HO fractions of the recycle solvent for PDU Run 5 are heavier than the corresponding fractions for Pilot Run 8. Therefore, the recycle oil during Pilot Run 8 was lighter than the recycle oil in PDU Run 5.

Comparison of elemental composition of the recycle solvent fractions indicates (see Table 14) that for both PDU and pilot runs, the HO distillate fraction contained more carbon, more nitrogen, and less hydrogen, when compared with the corresponding FB fractions. This trend is in agreement with the heavier nature of the OH distillate fractions. Similarly, the PDU recycle solvent fractions, i.e., FB and OH, contain more carbon and less hydrogen than the corresponding fractions of pilot plant recycle solvent. This trend also indicates that the recycle solvent during PDU Run 5 was heavier than that obtained during Pilot Run 8.⁽¹⁰⁾

Comparison of proton distribution of the hydroclone overflow distillates of Pilot Plant Run 8 and PDU Run 5 are shown in Table 15. It can be seen that the pilot plant distillates are less aromatic, more paraffinic, and have a lower ratio of cyclic/alkyl aliphatic protons. These parameters indicate that the pilot plant material should be a slightly poorer hydrogen donor solvent. Similar comparisons of the proton distributions for Pilot Plant Run 8 and PDU Run 5 fractionator bottoms show a similar trend, i.e., the

TABLE 15

PROTON DISTRIBUTIONS FOR PILOT PLANT AND
PDU RECYCLE SOLVENT COMPONENTS

	<u>Condensed Aromatics</u>	<u>Uncondensed Aromatics</u>	<u>Cyclic Aliphatic</u>	<u>Alkyl Aliphatic</u>
Pilot Plant Run 8 (FB)	17.3	7.2	31.4	44.1
Pilot Plant Run 8 (HO)	24.5	7.0	31.6	37.0
PDU Run 5 (FB)*	19.6	10.5	30.4	39.5
PDU Run 5 (HO)	26.7	11.1	29.2	33.0

*From the clean oil tank.

average PDU sample is more aromatic, less paraffinic, and has a high cyclic/alkyl aliphatic proton ratio (see Table 15).

The ideal process solvent should contain an optimum mix of aromatics, hydroaromatics, and paraffins. If the paraffin level exceeds this optimum as a result of overhydrogenation, precipitation of coal extract occurs, often resulting in coking and plugging. The paraffin content of the pilot plant recycle solvent, even though higher than the PDU recycle solvent, seems to be within the tolerable limits, because no unusual coking and plugging problems are reported for Pilot Plant Run 8.

Solvents with high aromatic content are shown to be better hydrogen donors than the predominantly paraffinic solvents.⁽¹¹⁾ For example, in laboratory micro-autoclave tests, the PDU Run 5 recycle solvent was more effective for the thermal dissolution of Indiana No. 5 coal than Pilot Plant Run 8 recycle solvent.⁽¹¹⁾ However, in

the H-Coal configuration, where the recycle solvent is in contact with a powerful catalyst, the importance of solvent quality is not as critical as it is in a noncatalytic thermal dissolution process, such as SRC I or II, and the first stage of the integrated two-stage liquefaction (ITSL) process. This is borne out by the fact that the pilot plant performance was comparable to that of the PDU.

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4.0 H-COAL COMMERCIAL PLANT DESIGN AND COST BASIS

As indicated earlier, the H-Coal liquefaction process has been a candidate for commercial demonstration. The first H-Coal commercial plant was proposed to be located in Breckinridge County, Kentucky. The initial H-Coal commercialization efforts were carried out under the Breckinridge project which was a cooperative agreement effort between the U.S. Department of Energy and private sector participants, consisting of Ashland Synthetic Fuels, Inc. (ASFI) and Airco Energy Company, Inc. (AECI).⁽¹⁾ Under this cooperative agreement, preliminary design (or Phase 0 design) cost estimate and plans for construction and operation of a first H-Coal commercial plant were developed.⁽¹²⁾ In this section, the technical basis used for developing the Phase 0 design and projecting the economics of the H-Coal commercial plant is presented.

4.1 Description of Proposed H-Coal Commercial Plant

The proposed H-Coal commercial plant is a fully-integrated grassroots facility. This includes provisions for on-site beneficiation and handling of run-of-mine (ROM) coal, conversion of coal into liquid and gaseous hydrocarbons, separation and purification of the products, stabilization and reforming of the naphtha, recovery of byproduct sulfur, ammonia, and phenols, production of process hydrogen by gasification of vacuum bottoms, and plant steam and utility production by on-site, coal-fired boilers. Additionally, wastewater treatment facilities, air

emission controls, and solid waste disposal are integral parts of the process design. Table 16 lists the commercial plant areas by name and number.

A detailed overall block flow diagram of the plant is shown in Figure 5.⁽¹³⁾ In the Phase 0 design, 23,582 TPD of ROM coal (dry basis) are received by the plant. Of this, 23,404 tons/day are sent to the coal beneficiation plant for ash and sulfur reduction. The washed coal (16,500 TPD) is then sent to the coal drying and pulverizing section to be prepared before slurrying and hydroliquefaction. The middlings (1,451 TPD) are sent to the steam plant. The tailings (5,453 TPD) are disposed of in landfills as a solid waste material.

In the H-Coal slurry preparation section of the plant, the total feed coal of 16,500 TPD is handled in seven operating trains with one spare. The capacity per train is 2,400 TPD. The slurried coal is pumped to reactor pressure and preheated in preheaters fired with plant-derived gas before entering the H-Coal reactor vessels. The eight trains are identical and consist of the following steps: slurry pumping, preheating, reaction and phase separation, and cooling. Plant 4 (H-Coal Primary Separation) consists of seven operating trains and one spare. The feed to Plant 4 is the reactor effluent vapor phase and a reactant effluent, liquefied product slurry phase. The vapor phase stream contains hydrogen, light hydrocarbons, steam, H_2S , NH_3 , and CO_2 . After partial cooling and water wash steps, this vapor is recycled back to the reactor via

TABLE 16

PLANT DESIGNATION FOR THE PHASE O BRECKINRIDGE PLANT

<u>Plant Number</u>	<u>Plant Area</u>
1	Coal Drying and Pulverizing
2	Coal Slurry Preparation
3	H-Coal Preheating and Reaction
4	H-Coal Primary Separation
5	H-Coal Recycle Slurry Preparation
6	H-Coal Recycle Hydrogen Compression
7	Gas Plant
8	Cryogenic Hydrogen Purification
9	Sour Water Treating*
10	Sulfur Plant*
12	Gasification and Purification
15	Oxygen Plant
17	H-Coal Distillate Separation
18	Naphtha Hydrotreating and Reforming
19	Flare System
20	Tankage
21	Interconnecting Piping
22	River Facilities
23	Rail, Truck, Pipeline
26	Coal Receiving and Storage
27	Coal Washing (Optional)
30	Electrical Distribution
31	Steam Generation and Boiler Feedwater Treating
32	Water Systems: Raw, Potable, Cooling
33	Fire Systems
34	Sewers and Wastewater Treatment*
35	Flue Gas Desulfurization
36	Instrument and Plant Air Systems
37	Communication Systems
38	Inert Gas Systems
39	Purge and Flush Oil Systems
40	Site Development Roads
41	Buildings
42	Solid Waste Management
43	Settling Ponds
44	Landfill*

*Plants using the proposed control technologies

Source: Reference 13

FIGURE 5
OVERALL BLOCK FLOW DIAGRAM

Plant 6. The liquefied product slurry is let down in pressure to an intermediate pressure flash vessel where H_2 and light gases are flashed off and routed to the gas plant (Plant 7). Further pressure letdown occurs in a low-pressure flash vessel where other light gases are removed and hydrocarbon liquids are recovered and sent to Plant 17 (distillate separation section). The remaining liquid slurry is fed to the hydroclones in the H-Coal recycle slurry preparation (Plant 5). At all stages of pressure letdown, wash water is injected to dissolve ammonium bisulfide, chloride, and bicarbonate salts which form as the reactor effluent stream is cooled. Water injection eliminates any potential plugging problems in lines and exchangers by dissolving these inorganic salts as they are formed.

Plant 5, H-Coal recycle slurry preparation, consists of two equal size operating trains which process the solids-containing product oil stream leaving the low-pressure letdown section of Plant 4. In Plant 5, the solids-containing oil slurry is first partially concentrated in a series of hydroclones. The hydroclone overhead containing about 14-percent solids is first flashed at atmospheric pressure to reduce light oil components of the slurry, before being returned to the coal slurry preparation plant (Plant 2).⁽¹⁴⁾ The hydroclone underflow is enriched with solids, containing approximately 24-percent solids, and is flashed to 15 psig, steam stripped, and vacuum-flashed for recovery of distillate hydrocarbon products. The residual vacuum bottoms stream

is a highly concentrated nonvolatile slurry primarily containing ash and IOM, and is sent to the gasification unit (Plant 12) for production of process hydrogen.

The various solids-free distillate streams separated from the H-Coal reactor effluent are combined in the distillate separation unit (Plant 17). These streams are fed to a fractionator feed heater and fractionation column where they are divided into light gases, a naphtha fraction, a middle distillate, and heavy oil. Plant-derived fuel gases are used as fuel to fire the fractionator feed heaters. Plant 17 consists of two operating trains. Its key processing objective is to maximize recovery of naphtha grade material (140-360°F boiling range). The middle distillate product consists of 360-650°F boiling range material. To improve the naphtha recovery, the fractionator middle distillate side-draw liquid is steam stripped. The heavy distillate product is a material with an initial boiling point (IBP) of 500°F and a minimum 140°F flash point. The heavy distillate product, in excess of the coal slurry preparation requirement, flows to storage. The middle distillate is pumped to storage and the heavy naphtha is delivered directly to the naphtha hydrotreating and reforming unit (Plant 18). All overhead vapours are sent to the gas plant (Plant 7).

Plant 7 consists of a single operating train, where feed streams from a number of sources are compressed, treated, and fractionated into intermediate and final products. The major

sections include gas compression, acid gas scrubbing using a di-ethanolamine (DEA) unit, product fractionation, liquid and gas drying, LPG sweetening, product naphtha sweetening, and refrigeration. The feed streams to Plant 7 originate from H-Coal primary separation, H-Coal distillate separation, naphtha hydrotreating and reforming, and cryogenic hydrogen purification. The products from Plant 7 are an ethane-rich fuel gas, LPG propane and butane, heavy naphtha, light straight run (LSR) gasoline, recycle hydrogen, acid gas, and sour water.

Plant 8, the cryogenic hydrogen purification plant, consists of a single operating train and operates in conjunction with the gas plant (Plant 7). It is used to purify a gaseous, hydrogen-rich stream received from the gas plant (Plant 7) and to deliver it to the make-up hydrogen gas compressor for process hydrogen requirements in Plant 3. The plant consists of a number of gas absorber beds followed by a cryogenic section, which is an insulated cold box. A sharp hydrogen separation is possible because of its extremely low liquefaction temperature. The unit produces the purified hydrogen and three byproduct streams. The first is principally ethane and heavier hydrocarbons; the next is pipeline fuel gas; and the third is a relatively inert waste gas.

The remainder of the make-up process hydrogen requirement is provided by Plant 12, the gasification and purification plant. This plant consists of four operating trains and one spare and includes the following units: Texaco gasification, including syngas

quenching and scrubbing; shift conversion and shifted gas cooling; shifted gas purification; hydrogen compression; and unshifted gas cooling and purification. The vacuum tower bottoms produced in the H-Coal recycle slurry preparation unit (Plant 5) are charged to Plant 12 and gasified in a series of five Texaco gasifiers (four to handle the vacuum bottoms and one spare). After gas quenching and scrubbing, the syngas stream is split, with the majority of the gas being sent to shift. The shifted gas is then cooled and passed to a selective Selexol unit for removal of CO_2 and H_2S . The shifted gas, which is essentially hydrogen, is compressed and sent to Plant 3 for reaction. The unshifted gas, i.e., that in excess of process hydrogen requirement, is cooled, purified in a Selexol unit, and then used as a medium-Btu plant fuel gas.

The Texaco gasification process produces a molten slag which is quenched in the bottom of the gasifier and removed through lockhoppers. The high-temperature operation of entrained gasifiers produces a syngas essentially free of hydrocarbons other than a small amount of methane (less than one percent). The acid gas removal (AGR) system used is a Selexol unit which selectively produces a hydrogen sulfide-rich stream suitable for Claus processing, and a carbon dioxide-rich stream containing only minor amounts of H_2S which can be vented to the atmosphere. Acid gases removed from the raw syngas are further processed to recover sulfur in a Claus sulfur recovery unit (Plant 10, which consists of two operating trains and one spare).

Oxygen required for the partial oxidation of the vacuum bottoms in the Texaco gasifiers is provided by conventional air separation units (Plant 15). Plant 15 consists of three equal-size operating trains. The steam generation plant (Plant 31) provides high-pressure, superheated steam at 900 psig and 750°F. The boilers are fired with the middlings coal from the coal washing plant. Each boiler is equipped with coal pulverizers, electrostatic precipitators, and induced draft fans. They also provide for collection and transfer of bottom ash and precipitator fly-ash. The plant design criteria for the boiler section calls for four 400,000-lbs/hr steam boilers to provide a normal steam production of 1,200,000 lbs/hr. The bulk of the high-pressure steam, about 47 percent, is used to run the air and oxygen compressors in Plant 15. The rest of the steam is used to drive two condensing turbines to produce about 37 MW of electric power. These turbines are equipped with a single extraction stage which provides 600 psig steam for driving the compressor in the gas plant (Plant 7) and also the pumps and fans in Plants 2, 3, 10, and 31.⁽¹³⁾ The boiler flue gas is cleaned in Wellman-Lord flue gas desulfurizers (Plant 35), reheated, and passed to 400-foot chimneys. The sulfur dioxide recovered in the FGD units (one FGD unit per boiler) is converted into elemental sulfur in the Claus sulfur plant (Plant 10), together with the other H₂S streams from the plant. The tail gas from the Claus unit is then incinerated in the boiler furnace.

Sour water streams originating from washing and quenching of liquid and gaseous products are collected and treated in Plant 9 (sour water treating). The sour water is first stripped to remove NH_3 and H_2S . The stripped water is then sent to a dephenolization unit for removal of phenols. The stripper overhead, containing the ammonia and acid gas, passes to an ammonia absorber unit (Phosam-W) where ammonia is absorbed and subsequently recovered as anhydrous ammonia. The H_2S -containing acid gas is then passed to the Claus unit for sulfur recovery. The dephenolized water then passes to the wastewater treatment section (Plant 34) for further processing. Plant 34 is a complex treatment facility capable of processing oily wastewater from the plant, sewage-containing waters, and process waters.

4.2 Phase 0 Commercial Plant Design Basis

The preliminary or Phase 0 design for the H-Coal commercial plant was based on the operational data collected during PDU Run No. 5. As indicated earlier, the catalyst activity reached the equilibrium value during the latter part of PDU Run 5. Therefore, the overall product yield structure obtained during Period 29 of PDU Run 5, in conjunction with HRI simulation of July 22, 1980, was used to develop flow quantities for different sections of the H-Coal commercial plant.^(1, 13) The product yield structure obtained during Period 29 of PDU Run 5 is given in Table 17.⁽⁴⁾ The total distillate, C_4^+ - 975°F, yield is about 47 percent of the dry

coal fed to the H-Coal reactor, which translates into approximately 3 Bbl of distillate per ton of dry coal.

4.2.1 Overall Product Yield Structure and Performance

The yield of C_4^+ distillate product expected from the H-Coal commercial plant is given in Table 18. The net output of C_4^+ - distillate product from the commercial plant is approximately 42.7 percent of the dry coal fed to the H-Coal reaction section. The total coal input to the H-Coal reaction section is about 16,500 TPD of dry coal. The net distillate output from the commercial plant is approximately 4.4 percentage point lower than the yield obtained during Period 29 of PDU Run 5 (see Table 17). This is a direct result of certain yield adjustments necessary for a commercial plant which are not required for a small PDU. For example, a small portion, about one percent of dry coal input, of the distillate oil, is used as a flush oil for slurry pumps used in the commercial plant. Also, a part of heavy naphtha, about one percent of dry coal input, is lost during the hydrotreating and reforming steps used for product upgrading in the commercial plant. A larger yield reduction, about 2.8 percent of dry coal input, has resulted because of using a lower end point for the distillate oil, 850°F as opposed to 975°F, produced in the commercial plant.⁽¹⁵⁾ After these adjustments are made, the overall distillate yield for the commercial plant agrees with the yield reported for PDU Run 5, Period 29.

TABLE 17

PRODUCT YIELD STRUCTURE
PDU RUN 5, PERIOD 29

<u>Run Designation</u>	<u>Lbs/100 lbs of Dry Coal</u>	<u>Bbl/ton of Dry Coal</u>
H ₂	-4.91	
H ₂ S	2.64	
NH ₃	1.08	
H ₂ O	6.67	
CO ₂	0.48	
C ₁	3.75	
C ₂	3.54	
C ₃	3.39	
C ₄	2.54	
C ₅ - 400°F	16.20	1.40
400 - 650°F	20.37	1.21
650 - 975°F	7.96	0.41
Residuum	19.00	
Unconverted Coal	5.78	
Ash	<u>11.51</u>	
TOTAL:	100.00	
Total Distillate, C ₄ - 975°F	47.1	3.01
Residuum, Unconverted Coal & Ash	36.3	

TABLE 18

C₄+ DISTILLATE PRODUCT YIELD STRUCTURE FOR PHASE 0 DESIGN

(Coal Input = 16,500 TPD D.B.)

	<u>Actual Rate lbs/hr</u>	<u>Percent of Dry Coal</u>	<u>Bbl/Ton of Dry Coal</u>	<u>Bbl/Day</u>	<u>Volume Percent</u>
C ₄	36,412	2.65	0.26	4,303	9.4
LSR*	36,712	2.67	0.22	3,742	8.2
Reformate	141,504	10.29	0.69	11,427	24.9
Middle Distillate (400 - 650°F)	291,195	21.17	1.27	21,050	} 57.5
Heavy Distillate (650 - 850°F)	<u>81,436</u>	<u>5.92</u>	<u>0.32</u>	<u>5,309</u>	
TOTAL C ₄ - 850°F Distillate Product	587,260	42.7	2.76	45,831	100.00
<u>Distillate Yield Adjustments:</u>					
(a) Flush Oil	14,001	1.0	0.055	--	
(b) Losses in Hydrotreating and Reforming	9,596	0.77	0.08	--	
(c) 85C - 975°F Fraction	<u>38,500</u>	<u>2.8</u>	<u>0.15</u>	<u>--</u>	
TOTAL C ₄ - 975°F Distillate Yield:					
Phase 0 Design	549,357	47.2	3.04		
PDU Run 5, Period 29	--	47.1	3.02		

* Light, straight-run gasoline.

For Phase 0 commercial design, approximately 57 percent (by volume) of the distillate product is middle and heavy distillate fraction; the bulk of it is middle distillate ($400^{\circ}\text{F} - 650^{\circ}\text{F}$) fraction. The gasoline boiling range material accounts for about 33 percent (about 8-percent light, straight-run gasoline and 25-percent reformat) of the total distillate product. The net C_4 , or butane, output is approximately 9 percent of the total distillate product.

The $\text{C}_1 - \text{C}_3$ yield structure for the Phase 0 commercial design is given in Table 19. The yield $\text{C}_1 - \text{C}_3$ products are in good agreement with the PDU Run 5 yields. The total commercial H-Coal plant product and by-product outputs for the Phase 0 design are given in Table 20. The overall estimated thermal efficiency of the Phase 0 H-Coal commercial plant is given in Table 21. The estimated thermal efficiency is about 61 percent, which is the ratio of the higher heating value of all the products divided by the higher heating value of total coal input plus the electrical energy input.

The overall thermal efficiency, however, is not an explicit indicator of the performance of a direct liquefaction process with respect to the yield of C_4 -distillate products. The performance of a direct liquefaction process can be expressed in terms of its overall hydrogen utilization efficiency. The hydrogen utilization efficiency here refers to the percent of total hydrogen, both hydrogen in the feed coal and that consumed in the process, input

TABLE 19

C₁ - C₃ PRODUCT YIELD STRUCTURE FOR PHASE O DESIGN

	<u>Actual Rate</u>		<u>Percent of Dry Coal</u>
	<u>lbs/hr</u>	<u>MMSCFD</u>	
C ₁	50,211	27.92	3.65 (3.75)**
C ₂	46,329	14.08	3.36 (3.54)
C ₃	<u>48,067</u>	<u>6,524*</u>	<u>3.49 (3.39)</u>
Total C ₁ -C ₃	144,607		10.5 (10.68)

*Bbl/day.

**Values in the parentheses are for PDU Run 5, Period 29.

TABLE 20

ESTIMATED PLANT OUTPUTS FOR PHASE O H-COAL COMMERCIAL PLANT

		<u>Units</u>
TOTAL COAL INPUT:	23,582	TPSD (D.B.)
<u>Major Plant Outputs:</u>		
SNG	31.78	MMSCFD
Refinery rich gas* (C ₂)	14.15	MMSCFD
Propane Product	6,524	BPSD
Butane Product	4,303	BPSD
Gasoline Blending Stock**	15,163	BPSD
Middle Distillate Product (400-650°F)	21,050	BPSD
Heavy Distillate Product (650-850°F)	5,309	BPSD
<u>Major By-Products:</u>		
Ammonia	193	TPSD
Sulfur	534	LTPSD
Phenols***	101	TPSD

*Approximately 30 percent of refinery gas is required to supply the in-plant fuel needs.

**Light gasoline product, plus reformate.

***Phenols are bonded with the distillate oil

TABLE 21

ESTIMATED PLANT THERMAL OUTPUT FOR
PHASE O H-COAL COMMERCIAL PLANT

<u>Energy Input:</u>	10 ⁹ Btu/day*
Coal**	534.84
Electricity***	<u>13.92</u>
TOTAL INPUT	548.76
Energy Output:	
SNG	30.19
Refinery Rich Gas	19.72
Propane Product	24.92
Butane Product	18.50
Gasoline Blending Stock	83.39
Middle Distillate Oil	124.2
Heavy Distillate Oil	<u>32.4</u>
TOTAL OUTPUT:	333.32
<u>Overall Thermal Efficiency:</u>	61%

*HHV of coal end products.

**Blend of Illinois No. 6 and Kentucky 9/11 coals.

***Conversion factor for electricity is 3,413 Btu/kWh at the plant gate.

used in producing C_4 -distillate products. In order to estimate the hydrogen utilization efficiency, an overall hydrogen balance is developed for the Phase 0 design and it is presented in Table 22.

Approximately 50 percent of the total hydrogen input to the commercial plant comes from the coal feed. The rest of the hydrogen comes from the gasification plant. Of the total hydrogen input, approximately 50 percent is used in producing C_4 -distillate products. Thus, the hydrogen utilization efficiency of the process is about 50 percent. Approximately 23 percent of the total hydrogen input is consumed in the production of C_1 - C_3 hydrocarbons. Therefore, there is a significant potential for increasing the hydrogen utilization efficiency by suppressing the formation of light hydrocarbons (C_1 - C_3), thus increasing the overall yield of C_4 -distillate products.

4.3 Phase 0 Commercial Plant Cost Basis

As a part of the DOE/Industry cooperative agreement mentioned earlier, detailed cost estimates were developed by Bechtel, Inc., to project economics for the construction and operation of a first-of-kind H-Coal commercial plant.⁽¹²⁾ The estimated plant capital and operating costs and the associated economic analyses are presented in this section.

4.3.1 Total Capital Cost Estimates

The total capital cost summary for the Phase 0 H-Coal commercial plant is given in Table 23. The capital cost data is extracted from the detailed cost estimates developed by Bechtel,

TABLE 22

OVERALL HYDROGEN BALANCE FOR PHASE O DESIGN

<u>Input</u>	<u>Actual Rate Lbs of H₂/Hour</u>	<u>Lbs of H₂/100 Lbs of Dry Coal</u>	<u>Percent</u>
Coal Feed	68,200	4.96	50.20
Hydrogen Consumed in the Process	<u>67,512</u>	<u>4.91</u>	<u>49.80</u>
Total H ₂ Input	135,712	9.87	100.
<u>Output</u>			
C ₁	12,553	0.92	9.4
C ₂	9,266	0.67	6.8
C ₃	8,787	0.64	6.5
C ₄	6,314	0.46	4.6
Reformate (a)	22,437	1.63	16.50
H ₂ Produced in the Reformer	1,957	0.14	1.4
Middle Distillate Oil	28,246	2.05	20.80
Heavy Distillate Oil	5,790	0.42	4.20
Flush Oil	955	0.07	0.70
975°F Material Left in the Vacuum Bottoms	2,737	0.20	2.0
Vacuum Bottoms (b)	19,800	1.44	14.60
H ₂ O, NH ₃ , H ₂ S, Fuel Gas, (c) and Losses	<u>16,870</u>	<u>1.23</u>	<u>12.50</u>
Total H ₂ Output	135,712	9.87	100.

22.7

50.2

(a) Reformate also includes light straight-run gasoline product.

(b) Residuum, IOM, and ash.

(c) Refinery-rich gas.

TABLE 23

CAPITAL COST SUMMARY FOR THE PROPOSED H-COAL COMMERCIAL PLANT

		MM \$ (January 1981)	Percent of Total Capital Requirement
I. <u>Major Process Plant Areas</u>			
<u>Plant Number(s)</u>	<u>Area Name</u>		
2, 3, 4, 5, 6	Liquefaction Plant Area	594.00	18.2
12, 15	Oxygen and Hydrogen Plants	323.00	9.9
7, 8, 17, 18	Product Separation, Refining, and Upgrading	143.00	4.5
1, 26, 27	Coal Handling	182.00	5.6
	Sub-total Process Plant Areas:	1,242.00	38.2
II. <u>Environmental Control Plant Areas</u>			
<u>Plant Number(s)</u>	<u>Area Name</u>		
9, 34	Sour Water/Waste Water Treatment Plants	136.00	4.2
10, 35	Sulfur and Stack Gas Scrubbing Plants	50.00	1.5
42, 43	Solid Waste Management, Landfill	29.00	0.9
	Sub-total Environmental Control Plants:	215.00	6.6
III. <u>Utility Plant Areas</u>			
<u>Plant Number(s)</u>	<u>Area Name</u>		
31	Steam Generation	188.00	5.9
30	Electrical Distribution	68.00	2.0
32	Water Systems: Raw, Potable, Cooling	34.00	1.0
36, 38, 39	Plant Air, Purge Gas, and Flush Oil Systems	12.00	0.4
	Sub-total Utility Plant Areas:	302.00	9.3

TABLE 23 (Concluded)

		MM \$ (January 1981)	Percent of Total Capital Requirement
IV. <u>Product Handling & Plant Support Facilities</u>			
<u>Plant Number(s)</u>	<u>Area Name</u>		
20, 21, 22, 23	Product Transfer, Storage, and Distribution Systems	118.00	3.6
19, 33, 37	Plant Fire, Flare, and Telecommunication Systems	26.00	0.8
40, 41	Site Development, Roads, and Buildings	<u>86.00</u>	<u>2.6</u>
	Sub-total Product Handling & Plant Support Facilities	<u>230.00</u>	<u>7.0</u>
TOTAL Direct Costs of All Plant Areas		1,989.00	61.1
V. <u>Other Capital Cost Components</u>			
	Initial Catalyst and Chemicals	29.00	0.9
	Field Indirects	400.00	12.3
	Engineering and Fee	292.00	8.9
	Miscellaneous	<u>50.00</u>	<u>1.5</u>
	Sub-total Plant Costs	2,760.00	84.70
	Contingency	<u>407.00</u>	<u>12.50</u>
TOTAL Plant Costs		3,167.00	97.20
	Working Capital	<u>91.00</u>	<u>2.8</u>
TOTAL CAPITAL REQUIREMENTS		<u><u>3,258.00</u></u>	<u><u>100.00</u></u>

Inc., under the Breckinridge Project Initial Effort.⁽¹²⁾ The subtotal plant costs are reported for major plant areas such as process plant areas, environmental control plant areas, utility plant areas, and product handling and support facilities. Other capital cost components include items such as initial catalyst and chemicals costs, field indirects, engineering and fee, etc. A contingency is added to the subtotal plant costs to arrive at total plant costs. Working capital is added to the total plant costs to arrive at total capital requirements. The assumptions used by Bechtel, Inc., for estimating the different components of the total capital requirements are briefly described below. However, detailed descriptions of specific cost estimating procedures can be found in the Breckinridge Project Initial Effort reports, or the "Blue books."⁽¹²⁾

The total capital requirement for the Phase U H-Coal commercial plant is 3,258 million dollars (January 1981 dollars). The total direct costs of all plant areas is approximately 61 percent of the total capital requirement. The direct costs of each plant area include the total installed cost of individual plants within that area. The installed cost of individual plants generally includes such items as field indirects and engineering and fee. However, for the sake of presentation, these two items are included under other capital cost components. The installed costs of individual plants was estimated by Bechtel, Inc. by using a factored cost estimating procedure based on the "bare equipment" costs⁽¹²⁾ developed from

Bechtel's in-house data and vendor estimates. Preliminary engineering design data were developed, particularly for first-of-a-kind plants, to estimate the bare equipment costs. For conventional plants, Bechtel's in-house data and vendor estimates were used to develop bare equipment costs. The costs for items such as foundations, piping, insulation, process buildings, instrumentation, etc., were estimated by taking a fixed percentage of the bare equipment costs. Additional factors were used for estimating the costs of items such as field indirects, engineering, home office allowances, and fee.⁽¹²⁾ The factors used for estimating these costs were developed, particularly for conventional plants, from historical data compiled by Bechtel on similar plants. However, for first-of-a-kind plants (i.e., H-Goal liquefaction section), the factors were developed primarily based on good engineering judgement of Ashland and Bechtel personnel.

The contingency allowance was estimated by using a risk analysis program to determine the overrun probability based on a Monte Carlo random number simulation.⁽¹²⁾ The results of the risk analysis are given in Figure 6. For a 50:50 probability of overrun on the project, the magnitude of the contingency allowance is estimated to be 15 percent of subtotal plant costs. With this contingency allowance, Bechtel/Ashland believes that the accuracy of the capital estimate ranged from +19 to -17 percent.⁽¹²⁾ Working capital consists of a reasonable allowance which covers expenses such as cash necessary to meet payroll obligations, accounts

Source: Reference 12

FIGURE 6
RESULTS OF RISK ANALYSIS

receivable, and raw material inventory. The coal, catalyst, and chemicals inventory costs are estimated at 60 days of storage. The payroll obligations and accounts receivable are estimated at 10 percent of annual operating costs.⁽¹⁷⁾

4.3.2 Annual Operating Costs

The annual operating costs, as estimated by Bechtel, Inc., are presented in Table 24.⁽¹⁶⁾ The operating costs are broken down into three major categories: variable costs, which include items such as raw materials, catalyst, chemicals, and utilities; semi-variable costs, which include labor and labor-related costs, and fixed or capital-related charges.

The total annual estimated operating costs are about \$452 million dollars. The largest component of the total annual operating costs is the cost of run-of-mine coal consumed at the plant. The coal costs represent approximately 54 percent of the total annual operating costs. The purchased electric power cost constitutes approximately 10 percent of the total annual operating costs. The estimation of this variable operating cost component was based on the detailed material and energy balance calculations.

The labor-related expenses amount to approximately 9 percent of the total annual operating costs. In order to estimate the labor-related costs, a detailed organization chart for the management and operation of the plant was developed. The breakdown of the total personnel requirement is given in Table 25.⁽¹⁶⁾ The labor costs include wages and annual salaries of plant operation and

TABLE 24

ANNUAL OPERATING COST
(87% Stream Factor)

	January 1981 Dollars
<u>I. Feedstock and Utilities Costs</u>	
Coal - \$29/ton, 26,497 TPSD*	242,280,600
Purchased Electric Power 170,000 kWh/hr at 3.654/kWh	46,954,500
Annual Catalyst and Chemicals	<u>34,380,000</u>
Subtotal:	323,615,100
<u>II. Labor Costs</u>	
Direct Labor and Supervision	13,265,300
Maintenance Labor and Supervision	14,965,900
Contract Maintenance	<u>11,343,100</u>
Subtotal:	39,574,300
<u>III. Supplies and Materials</u>	
Rolling Stock Fuel and Tires	3,045,000
Operating Supplies	7,241,500
Maintenance Materials	<u>21,967,700</u>
Subtotal:	32,254,200
<u>IV. Other Costs</u>	
Indirects** and G&A	8,782,500
Local Taxes and Insurance	<u>47,511,500</u>
TOTAL Annual Operating Cost	451,737,600

* As Received run-of-mine coal.

** Indirect costs include items such as laboratory, safety, health, and security, etc. related expenses.

TABLE 25

PERSONNEL REQUIREMENT FOR THE PROPOSED H-COAL COMMERCIAL PLANT

	<u>Number of Persons</u>
Vice President and General Manager's Office	2
Laboratory	62
Operations	425
Engineering	58
Environmental, Health, Safety, and Security	54
Administration	121
Maintenance (In-house and Contract)	<u>823</u>
TOTAL	1,545

maintenance personnel. The wages for operations and in-house maintenance personnel were based on the H-Coal pilot plant, Catlettsburg, Kentucky, personnel wage scales in force as of January 1981.⁽¹⁶⁾ The wages for the contract maintenance personnel were estimated based on Ashland Oil Company's experience at a comparable size refinery in Louisville, Kentucky. The annual salaries for the salaried personnel were based on Ashland's salary structure.

The costs of operating supplies are estimated at 15 percent of the total maintenance cost, whereas the maintenance materials are estimated at 100 percent of maintenance salaries and wages and fringe benefits.⁽¹⁶⁾ Rolling stock fuel and tires expenses are estimated based on surface mining experience. Local taxes and insurance amount to approximately 10.5 percent of the total annual operating costs. They are estimated by adjusting the prevailing rates in the Breckinridge County area.

4.3.3 Economic Analysis

The capital cost and annual operating cost estimates presented earlier were used by Bechtel Petroleum, Inc. to conduct a detailed economic analysis. This section will present a summary of assumptions used in conducting the economic analysis and will highlight the results of such analysis. Detailed information on the methodology and assumptions is given in the Ashland/Bechtel Report.⁽¹⁷⁾

The economic performance of the project was estimated based on the projected product values developed by the PACE Company,

Engineers and Consultants, Inc.⁽¹⁷⁾ The market values for the products and by-products produced from the proposed H-Coal commercial plant were projected under different economic scenarios. Only the results of the base case economic scenario are highlighted in this report.

4.3.3.1 Highlights of Base Case Economic Scenario. The detailed methodology used in forecasting the market price and share for the products/by-products from the first H-Coal commercial plant under the base case economic scenario is given elsewhere.⁽¹⁷⁾

Briefly, the base case economic scenario assumes that the world economic growth, including the United States, will be significantly less than the historical trend established during the past twenty years.⁽¹⁷⁾ In addition, under this scenario, major disruptions in free world crude oil supplies are not anticipated. Therefore, it is projected that prices in real terms for the crude oil by the end of this century will be slightly higher than the peak prices of 1981. The projected prices for the crude oil, both in current and constant 1981 dollars, are given in Table 26. Under the base case economic scenario, it is projected that the crude oil prices will rise to \$41.5/bbl (1981 dollars) in 1990, and stay at this level through the end of the century.

TABLE 26

PROJECTED PRICES FOR CRUDE OIL UNDER BASE CASE ECONOMIC SCENARIO

	Dollars per Barrel		
	<u>1985</u>	<u>1990</u>	<u>2000</u>
In Current Dollars	47.90	76.03	138.65
In Constant 1981 Dollars	35.34	41.50	41.50

With the above established base for average U. S. crude oil prices, a detailed supply/demand forecast is developed for the various petroleum-refined products from Gulf Coast refineries.⁽¹⁷⁾ The market prices for refined petroleum products were then adjusted for quality and geographical location. The adjusted product values for the H-Coal commercial plant products are developed over the life of the project. The projected product values for the year 1991, the year in which the commercial plant is expected to achieve full capacity, are presented in Table 27. The product values are presented both in constant 1981 and current dollars. These projected product values formed the basis for estimating the economic performance of the proposed H-Coal commercial plant in the following section.

The economic performance of the proposed H-Coal commercial plant is estimated both by a discounted cash flow, rate of return on equity (DCF-ROE), and a required product price analysis. In the former, the expected product market prices are used to calculate a discounted cash flow rate of return on the equity portion of total

project costs. In the latter, for a desired DCF-ROE, the required product price is calculated. The results of both analyses follow.

TABLE 27

BRECKINRIDGE PROJECT PRODUCT VALUES (YEAR 1991)

<u>Product</u>	<u>Unit</u>	<u>1981 Dollars</u>	<u>Current Dollars</u>
SNG	MCF	6.87	14.35
Propane	Gal	0.845 (35.5)*	1.705 (74.1)
Butane	Gal	0.905 (38.01)	1.891 (79.4)
Light Gasoline	Gal	0.994 (41.74)	2.07 (86.90)
Reformate	Gal	1.38 (57.96)	2.89 (121.4)
Distillate	Gal	1.04 (43.68)	2.173 (91.3)
Sulfur	L.T.	222	464
Ammonia	S.T.	260	543

*The numbers in parentheses are in \$/Bbl.

4.3.3.2 Results of Base Case Economic Analysis. The net outputs of products/by-products from the proposed H-Coal commercial plant are given below:

● Propane	5,676 BPCD*
● Mixed Butanes	3,744 BPCD
● Light Naphtha	3,255 BPCD
● Reformate	9,936 BPCD
● Distillate Oil	22,932 BPCD
● By-products	
SNG	39.9 MMSCFD**
Sulfur	464.6 LTCD
Ammonia	168 TPCD

*Barrels per calendar day.

**Includes refinery-rich gas (C₂).

The net production rate of the products is obtained by multiplying the plant name plate production capacity with stream factor of 0.87. The propane, SNG, sulfur, and ammonia produced will meet the current commercial product specifications.⁽¹⁷⁾ The C₄-distillate products are to be marketed as blending components. The C₄ fraction produced at the H-Coal plant consists of mixed isomers with a high percentage (about 74 percent) of normal butane. The rest of the C₄ fraction will contain iso-butanes, butylenes, and small amounts of propane and pentane.⁽¹⁷⁾

The light naphtha and reformate will be marketed as gasoline blending stocks. The light naphtha fraction will be comparable to a C₅-140°F petroleum cut with an R+M/2 octane of 62 and a Reid vapor pressure of 9. The reformate fraction is expected to have a high octane rating of 98.5 and Reid vapor pressure of 5.⁽¹⁷⁾ The distillate fuel oil will be utilized for its thermal value alone, or it may be considered as a cutter stock for refinery-produced heavy fuel oils.

The compatibility of H-Coal distillate fuel oils with their counterpart petroleum-based oils should be confirmed before they can be marketed as blending stock. Except for the light, straight-run (LSR) gasoline and reformate fractions, the other H-Coal distillate oil fractions are expected to have significantly high nitrogen and oxygen content. However, as indicated earlier, for the purposes of this economic analysis, certain product value adjustments based on quality differentials are made.

TABLE 28

BASE CASE ECONOMIC ASSUMPTIONS

Products Price Forecase:	Based on crude oil valued at \$35/bbl (1981\$) increasing to \$41.50/bbl by 1990, and remaining at that value through the life of the project
Inflation:	Average 8.5%/year through construction, and 6.0% during operating life of the project
Tax Credits:	15% of total fixed capital
Tax Depreciation:	Project life, 15 years; sum of years digits followed by double declining balance
Plant Stream Factors:	20% - 1988, 50% - 1989, 80% - 1990, 86.4% - 1991 through project life
Type of Financing:	100% Equity 48/52% Debt/Equity*

*Corresponds to a \$3 billion loan limit set by the U.S. Synthetic Fuels Corporation (SFC).

The major assumptions used in the base case economic analysis are given in Table 28. The total capital requirements for the H-Coal commercial plant are given in Table 29. The capital requirements are presented both in constant 1981 dollars and in "as spent", or current dollars. The constant 1981 dollars represent the overnight construction costs, which do not take into account the actual capital expenditures spread over the design/engineering, procurement, construction, commissioning, and startup periods. Therefore, the constant 1981 dollars, or overnight construction costs, do not give the actual magnitude of funds required for financing the project. The "as spent" capital requirements, however, are a realistic estimate of total funds required to implement the proposed H-Coal commercial plant. As shown in Table 29, the total cumulative capital requirements are approximately six billion dollars. It is assumed that these funds will be spent between 1981 and 1988, a period of seven years, during which the detailed design/engineering, procurement, construction, and plant startup activities will take place. It is expected that the plant will achieve full rated capacity in the beginning of 1991.

TABLE 29

BASE CASE TOTAL CAPITAL REQUIREMENTS (1981 - 1988)

	<u>1981 Dollars</u>	MM\$	<u>Current Dollars</u>
Equipment and Construction	3,167.00		4,491.00
Working Capital	91.00		36.00**
Interest during Construction	--		1,407.00
TOTAL	<u>3,258.00*</u>		<u>5,934.00***</u>

*Overnight construction costs.

**Working capital required in 1988.

***Overall funding requirement.

The key deflators used for converting the constant 1981 dollars to current dollars are given in Table 30. The base case economic analyses are carried out both for 100-percent equity financing and also for debt financing. For the debt financing, it is assumed that the debt portion, a maximum of three billion dollars, will be guaranteed by the U.S. Synthetic Fuels Corporation. The interest rate on the debt fraction is estimated to be 10.8 percent.⁽¹⁷⁾

The product pricing projections, developed as a part of the detailed market analysis mentioned earlier, have been used to estimate the annual revenues generated from the sale of the products over the life of the project. An example of base case product prices and annual revenues is given in Table 31. The product prices and revenues are given in current dollars for 1991, the first year in which the plant achieves full production rate. The composite expected market price for the liquid products (C_3 -850°F) is about \$94.4/Bbl in 1991 dollars. The results of the discounted cash flow return on equity (DCF-ROE) analysis for both equity and debt financing are given in Table 32. With 100-percent equity financing, the DCF-ROE is about 12 percent, which is a low rate of return to attract private investment. However, the assumed debt financing improves the economic viability of the project considerably, as it yields a DCF-ROE of about 21 percent, which is more acceptable, and in most cases, above the corporate minimum cutoff rates. (It is assumed that a typical corporate minimum cutoff rate is about 15 percent.)

TABLE 30

OTHER KEY BASE CASE ASSUMPTIONS

United States Gross National Product Deflator Forecast

<u>Year</u>	<u>Annual % Change</u>	<u>Index (1981 = 1.0)</u>	<u>Year</u>	<u>Annual % Change</u>	<u>Index (1981 = 1.0)</u>
1981	9.8	1.000	1987	6.7	1.623
1982	9.7	1.098	1988	6.5	1.732
1983	8.7	1.205	1989	6.5	1.845
1984	7.6	1.309	1990	6.3	1.965
1985	7.5	1.409	2000	5.8	3.453
1986	7.2	1.514			

A required product price analysis was also conducted to determine the level of composite prices of all liquid products that would yield a desired rate of return on the equity portion of the total project funds. All the economic assumptions used in the DCF-ROE analysis were also used in carrying out the required product price analysis. The required product prices to yield a 15-percent return on equity, both for 100 percent equity and debt financing, are given in Table 33. A 15-percent rate of return is selected because it represents a minimum hurdle or cutoff rate commonly used in private sector capital project evaluation. As shown in Table 33, for the 100-percent equity financing case, the required prices are significantly higher than the projected or expected market prices. Once again, the debt financing has significantly improved the commercial viability of the project, because the required prices are approximately 28 percent lower than the market price.

TABLE 31

EASE CASE PRODUCT PRICES AND REVENUES
FOR PHASE O H-COAL COMMERCIAL PLANT

(Current Dollars)

	Average 1991 ^(a) Rate per Calendar Day	Units	Price per Unit (\$)	Annual Revenue (MM \$)
<u>Liquid Products</u>				
Distillate Oil	22,774	Bbl	91.14	757.60
Reformate	9,873	Bbl	121.38	437.40
Propane	5,637	Bbl	74.13	152.50
Butane	3,718	Bbl	79.42	107.70
Light Gasoline	<u>3,233</u>	Bbl	87.19	<u>102.80</u>
	45,235	Bbl	94.4 ^(b) (45.2) ^(d)	1,558.00
<u>By-Products</u>				
Pipeline Gas	39,959 ^(c)	MSCF	13.76	200.70
Sulfur	461.4	Long Tons	464	78.1
Ammonia	168.2	Tons	543	<u>33.3</u>
TOTAL REVENUE				1,870.1 (895.86) ^(d)

(a) 1991 represents the first year at full production rate.

(b) Composite price for C₃+ - 850°F liquid product.

(c) Includes about 30 percent refinery-rich gas.

(d) The numbers in parentheses are in constant 1981 dollars.

TABLE 32

RESULTS OF DISCOUNTED CASH FLOW
RETURN ON EQUITY (DCF-ROE) ANALYSIS
(Current Dollars)

<u>Type of Financing</u>	<u>DCF-ROE</u>
100-percent equity	12.2
Debt financing to \$3 billion SFC loan limit	20.8

TABLE 33

COMPARISON OF EXPECTED MARKET AND REQUIRED PRICES
FOR C₃-LIQUID COMPOSITE PRODUCT

<u>Operating Years:</u>	<u>1988</u>	<u>1993</u>	<u>1998</u>	<u>2003</u>
<u>Product Price:</u>	<u>\$/Bbl**</u>			
Expected Market Prices*	73.59	104.62	140.01	187.36
Required Prices				
15% DCF-ROE (100% equity financing)	89.78	127.64	170.81	228.58
15% DCF-ROE (Debt financing)***	57.4	81.6	109.21	146.14

*Based on base case market forecasts.

**Current dollars.

***Financing to \$3.0 billion SFC loan limit.

A sensitivity analysis was also conducted by Bechtel, Inc. to determine the impact of certain significant variables on project economics.⁽¹⁷⁾ The variables considered include:

- Product pricing
- Equipment and construction costs
- Coal costs
- Stream factor

The results of the sensitivity analysis are presented in Table 34. The sensitivity analysis for the most part was conducted only for the 100-percent equity financing case. Among all the variables tested, only the product pricing sensitivity has relatively significant impact on the DCF-ROE. For example, a 24-percent increase in the product price will increase the discounted cash flow return on equity (DCF-ROE) by about four percentage points. A 14-percent increase in the equipment and construction costs (or total plant costs) has decreased the DCF-ROE by one percentage point. However, the DCF-ROE is relatively more sensitive to changes in the plant costs for the debt financing case. For example, a 14-percent increase in the plant costs has decreased the DCF-ROE by four percentage points in the debt financing case. Thus, a DCF-ROE of 17 percent will be achieved as opposed to 21 percent in the base case. This rate of return is slightly over the typical non-risk adjusted corporate hurdle rate of 15 percent.⁽¹⁷⁾ Therefore, unless the total project cost in the detailed engineering phase is shown to be within the 50/50 probability estimates developed under Phase 0 efforts, the project may not be economically attractive in spite of the assumed debt leveraging. The DCF-ROE is not sensitive to changes in annual operating costs and plant stream factors. For example, a 5-percent increase in the annual

TABLE 34

RESULTS OF SENSITIVITY ANALYSIS (100% EQUITY FINANCING)

<u>Variable/Sensitivity</u>	<u>Change in DCF/ROE</u>
Liquid Product Prices:	
24% increase	4.2% increase
Equipment and Construction Costs:	
14% increase	1.0% decrease (4.1% decrease for debt financing)
Annual Operating Costs:	
5% increase	0.4% decrease
5% decrease	0.5% increase
Stream Factor:	
90%	0.4% increase
82%	0.4% decrease

operating costs, which corresponds to roughly 10-percent increase in coal costs, will decrease the DCF-ROE only by 0.4 percentage point. Similarly, increasing the plant stream factor from 87 percent to 90 percent will increase the DCF-ROE only by 0.4 percentage point. Also, a decrease in plant factor from 87 percent to 82 percent will reduce the DCF-ROE only by 0.4 percentage point.

5.0 IMPACT OF PILOT PLANT OPERATIONS

The operating results, obtained during Pilot Run 8, four-day material balance period, are incorporated in the Phase 0 H-Coal commercial plant design basis. The impacts, if any, of incorporating the pilot plant data on the projected technical and economic performance of the Phase 0 H-Coal commercial plant are assessed.

5.1 Modifications of Design Basis, Based on Pilot Plant Operations

As indicated in the earlier sections, the Phase 0 H-Coal commercial plant design was based on the results of PDU runs obtained during Period 29. Since that time, the H-Coal pilot plant in Catlettsburg, Kentucky has made several long-term, successful runs on the candidate coal feedstock. The product yields and distribution obtained during the successful pilot plant runs were, for the most part, similar to those obtained during PDU Run 5, except for the yield of C_3-C_4 fractions and the total hydrogen consumption. The detailed comparison of operating results from PDU Run 5, Period 29, and Pilot Plant Run 8, four-day material balance period, is presented in Section 3.4 of this report. Pilot Plant Run 8, four-day material balance period, represents operating conditions similar to those expected for the H-Coal commercial plant. Also during this period, steady-state conditions prevailed, and the overall material balance closure of about 98 percent was reached.

The overall product yield structure obtained during this period of Pilot Plant Run 8 is given in Table 35.⁽⁵⁾ This product yield

TABLE 35

PRODUCT YIELD STRUCTURE FOR PILOT RUN 8,
FOUR-DAY MATERIAL BALANCE PERIOD

<u>Output</u>	<u>Run Designation</u>	
	<u>Lbs/100 Lbs of Dry Coal</u>	<u>Bbl/ton of Dry Coal</u>
H ₂	-5.47	
H ₂ S	2.67	
NH ₃	0.65	
H ₂ O	6.67	
CO ₂	----	
C ₁	4.20	
C ₂	3.02	
C ₃	4.55	
C ₄	4.61	1.69
C ₅ - 400°F	17.8	1.69
400 - 650°F	16.46	0.99
650 - 975°F	8.81	0.46
Residuum	21.26	
Unconverted Coal	3.46	
Ash	11.31	
TOTAL:	100.00	
Total Distillate, C ₄ ⁺ -975°F	47.7	3.14
Residuum, Unconverted Coal and Ash	36.0	

structure and other pertinent operating results are used to modify the Phase 0 H-Coal commercial plant design.

5.2 Description of Modified Phase 0 Design

The overall configuration of the H-Coal commercial plant described in Section 4.1.1, will not change as a result of minor design modifications discussed in this section. The modifications are required to incorporate the operating results of the H-Coal pilot plant in the Phase 0 commercial plant design. The major modifications are:

- Revision of overall product yield structure
- Expansion of gas plant (Plant 7) to accommodate increased propane and butane production
- Expansion of gasifier plant (Plant 12) to process slightly higher quantities of vacuum bottoms
- Small increase in oxygen plant (Plant 5) capacity
- Expansion of steam plant (Plant 31) to meet the increased high-pressure steam demand of the oxygen plant

In addition to these modifications, minor modifications may also be required in the other plants. However, it is assumed that the minor modifications will not impact the technical and economic performance of the H-Coal commercial plant. The technical and economic impacts of the above-mentioned modifications are discussed below.

5.3 Impacts on Technical Performance

The yield structure of C_4^+ -distillate product for the modified Phase 0 commercial design is given in Table 36. The net yield of C_4 -850°F distillate fraction is about 39.6 percent of the dry coal fed to the H-Coal reactor. After the necessary yield adjustments, i.e., flush oil, losses in hydrotreating and reforming and 850-975°F fraction left in vacuum bottoms, the gross yield of C_4 -975°F fraction becomes about 48.3 percent of the dry coal feed. This gross yield is in good agreement with the yield reported for Pilot Run 8, four-day material balance period.

The yield structure for C_1 - C_3 compounds is given in Table 37. The modified Phase 0 design produces approximately

TABLE 36

C₄+ DISTILLATE PRODUCT YIELD STRUCTURE FOR REVISED PHASE C DESIGN

(Coal Input = 16,500 TPD D.B.)

	<u>Actual Rate</u> <u>lbs/hr</u>	<u>Percent of</u> <u>Dry Coal</u>	<u>Bbl/Ton of</u> <u>Dry Coal</u>	<u>Bbl/Day</u>
C ₄	66,579	4.84	0.47	7,872
LSR	27,781	2.01	0.17	2,832
Reformate	155,634	11.30	0.76	12,562
Distillate Oil* (400 - 850°)	<u>295,492</u>	<u>21.49</u>	<u>1.26</u>	<u>20,858</u>
TOTAL C ₄ - 350°F Distillate Product	545,486	39.65	2.66	44,124
<u>Distillate Yield Adjustments</u>				
(a) Flush Oil	14,001	1.0	0.055	--
(b) Losses in Hydrotreating and Reforming	10,554	0.77	0.03	--
(c) 975°F Fraction Left in Vacuum Bottoms	<u>94,869</u>	<u>6.9</u>	<u>0.37</u>	<u>--</u>
TOTAL Phase C Design	664,910	48.32	3.16	--
Pilot Plant Run 8 Four-day Material Balance	--	48.7	3.21	--

* Bulk of this oil is assumed to be middle distillate with boiling point range 400 - 650°F.

TABLE 37

C₁ - C₃ PRODUCT YIELD STRUCTURE FOR REVISED PHASE O DESIGN

	<u>Actual Rate</u>		<u>Percent of Dry Coal</u>
	<u>lbs/hr</u>	<u>MMSCFD</u>	
C ₁	58,850	33.9	4.28 (4.28)**
C ₂	42,350	12.8	3.08 (3.08)
C ₃	<u>60,753</u>	<u>8,246*</u>	<u>4.4 (4.6)</u>
	161,953	-	11.76

* Bbl/day.

**Values in the parenthesis are for Pilot Plant, Run 8; four-day material balance period.

26 percent more propane and 82 percent more butane, respectively, than the original Phase 0 design. Because of the higher yields of butane, the overall C_4 -850°F distillate fraction yield structure has significantly changed. For example, the volumetric yield of C_4 fraction has increased from 9 percent total C_4 -distillate yield for the Phase 0 design to approximately 18 percent for the modified Phase 0 design. Finally, the reformat yield for the modified Phase 0 design has increased by about four percentage points.

Also, in the modified Phase 0 design, approximately 12 percent more vacuum bottoms are processed in the gasifier or hydrogen plant (Plant 12). This increase is achieved by leaving more 850-975°F distillate material in the vacuum bottoms. The vacuum bottoms fraction was cut back, partly because of the pumping considerations, and partly due to the increased hydrogen consumption of the modified Phase 0 design. The hydrogen consumption of the modified Phase 0 design is approximately 11 percent higher than the Phase 0 design.

Because of the increased quantity of vacuum bottoms processed in Plant 12, the amount of O_2 and steam required in the Texaco gasifier will proportionately increase. Since the high-pressure steam used in the Texaco gasifier is produced in the waste heat recovery boilers of the shift conversion plant, an increase in hydrogen production capacity will proportionately increase the

availability of high-pressure steam for the gasifier. Thus there will be no net demand for the high-pressure steam from the gasifier plant.

The additional vacuum bottoms fed to the gasifier are equivalent to about 58,000 lbs. of carbon per hour. Since the oxygen to carbon weight ratio within the gasifier is maintained at about 1.08, an additional 63,000 lbs. per hour of pure oxygen will be required for the gasifier operation. Therefore, the oxygen plant capacity will increase from the normal operating rate of 4,528 tons per day to 5,285 tons per day, an increase of about 16 percent. However, the total rated capacity of the oxygen plant (Plant 15) is about 5,000 tons per day.⁽¹⁸⁾ Therefore, the actual expansion in the oxygen plant capacity will be around 6 percent.

The C_3^+ -liquid output from the gas plant (Plant 7) will increase from 23,300 BPSD for Phase 0 design to 27,200 BPSD for the modified Phase 0 design, an increase of about 16 percent.

Because of the required capacity expansions in the gas plant (Plant 7), the hydrogen plant (Plant 12), and the oxygen plant (Plant 15), the utility consumption of these plants will increase. For example, it is estimated that the electricity consumption in Plant 12 will increase from 48,252 kWh/hr to 54,166 kWh/hr, a net increase of about 12 percent. The bulk of this increased electric power consumption will be used for the increased power requirement of the recycle hydrogen compressor.

The oxygen plant (Plant 15) has three operating trains. The compressors, both air and O_2 of one train are driven by electric motors, whereas the compressors for the other two trains are driven by steam.⁽¹⁶⁾ For the modified plant, it is assumed that the additional O_2 capacity is equally distributed among these three operating trains. The incremental electric power consumption for the single O_2 train is estimated to be about 3,683 kWh/hr.

Plant 7, the gas plant, is a net exporter of electric power. Therefore, the Phase 0 design modifications will not result in additional power consumption.

The overall flow of high-pressure steam for the proposed H-Coal commercial plant is given in Figure 7.⁽¹³⁾ The steam plant (Plant 31) produces about 1,200,000 lbs/hr of 900-psig, 750°F steam. The bulk of the high-pressure steam, about 47 percent, is used to run the air and oxygen compressors in Plant 15.⁽¹³⁾ The rest of the steam is used to drive two condensing turbines to produce about 37 MW of electric power.⁽¹³⁾ These turbines are equipped with a single extraction stage which provides about 145,000 lbs/hr of 600 psig steam for driving the refrigerant compressor in Plant 7, and also the pumps and fans in Plants 2, 3, 10, and 31. Approximately 97,000 lbs/hr of extracted steam is consumed in Plant 7 for the Phase 0 design.

The minor capacity revisions in Plant 15 (the oxygen plant) and Plant 7 (gas plant) will involve the following incremental high- and intermediate-pressure steam demand:

Source: Reference 13.

FIGURE 7
HIGH PRESSURE STEAM FLOWS FOR THE PROPOSED
H-COAL COMMERCIAL PLANT

	<u>Incremental Demand</u>
High-pressure steam, 950 psig, 750°F	68,500 lbs/hr
Intermediate-pressure steam, 600 psig, 665°F	32,500 lbs/hr

For the modified Phase 0 design, it is assumed that the increased steam demand, both at the high- and intermediate-pressure levels, will be met by expanding the steam plant (Plant 31) capacity.

It is estimated that approximately 97,500 lbs/hr of additional 900-psig, 750°F steam will have to be produced in the steam plant in order to satisfy the incremental steam demand resulting from the minor capacity revisions in Plants 15 and 7. Approximately 70 percent of this additional steam is used in Plant 15, to provide the increased power requirements for both the air and oxygen compressors. The balance of the additional high-pressure steam will be used to increase the throttle rate of one of the condensing turbines. The increased throttle rate will provide the additional extraction steam required for the revised Plant 7, without reducing the net electric power output from that turbine.

The total commercial H-Coal plant product and by-product outputs based on the modified product yield structure are estimated in Table 38. As indicated earlier, the propane and butane production rate for the modified Phase 0 plant has increased by 1,722 and 3,569 Bbl/day, respectively. The reformato production has increased by 1,161 Bbl/day. The distillate oil production, however, has decreased

TABLE 38

ESTIMATED PLANT OUTPUTS FOR PHASE 0 AND MODIFIED PHASE 0
COMMERCIAL H-COAL PLANTS

	<u>Phase 0 Plant</u>	<u>Modified Phase 0 Plant</u>	<u>Units</u>
TOTAL COAL INPUT		23,582	TPSD (D.B.)
<u>Major Plant Outputs:</u>			
SNG	31.78	31.78	MMSCFD
Refinery rich gas* (C ₂)	14.15	14.15	MMSCFD
Propane Product	5,524	8,246	BPSD
Butane Product	4,303	7,872	BPSD
Gasoline Blending Stock**	15,163	15,393	BPSD
Middle Distillate Product (400-650°F)	21,050	20,858***	BPSD
Heavy Distillate Product (650-850°F)	5,309		BPSD
<u>Major By-products:</u>			
Ammonia	193	193	TPSD
Sulfur	534	534	LTPSD
Phenols	101	101	TPSD

*Approximately 30 percent of refinery gas is required to supply the in-plant fuel needs.

**Light gasoline product, plus reformate.

***Bulk (about 90 percent by volume) of this distillate oil is assumed to be middle distillate (400-650°F) fraction.

by about 5,500 Bbl/day. The decrease in distillate oil yield corresponds to the elimination of heavy distillate oil (650-850°F) as the net distillate oil product for the modified Phase 0 design. The heavy distillate oil fraction is used to cut back the vacuum bottoms fraction, and it is eventually gasified in Plant 12. The overall estimated thermal efficiency for the modified Phase 0 plant is given in Table 39. This also compares the efficiency of the Phase 0 design. The reduction in the production of relatively high energy content distillate oil fraction, i.e., heavy distillate oil, coupled with increased steam and electricity consumption, has resulted in a three-percentage point drop (58 percent as opposed to 61 percent) in the overall thermal efficiency for the modified Phase 0 design. However, while the overall thermal efficiency has decreased for the modified Phase 0 plant, the yield of readily marketable distillate products has significantly increased; therefore, the thermal efficiency loss will not penalize the overall performance of the commercial plant.

As indicated earlier, the hydrogen utilization efficiency is an additional indicator of the performance of a direct liquefaction process. The overall hydrogen balance for the modified Phase 0 commercial plant is presented in Table 40. As was the case with the Phase 0 commercial plant, approximately 50 percent of the input hydrogen to the liquefaction plant is used to produce C_4^+ -distillate liquid products. On the other hand, the amount of hydrogen used for producing C_1 - C_3 hydrocarbons has increased from

TABLE 39

OVERALL THERMAL EFFICIENCY OF PHASE 0 AND REVISED PHASE 0
COMMERCIAL H-COAL PLANTS

	<u>Phase 0</u>	<u>Revised Phase 0</u>
<u>Energy Input:</u>	10^9 Btu/day *	
Coal**	534.84	
Electricity***	<u>13.92</u>	
TOTAL INPUT:	548.76	
<u>Energy Output:</u>		
SNG	30.19	30.19
Refinery Rich Gas	19.72	19.72
Propane Product	24.92	31.49
Butane Product	18.50	33.84
Gasoline Blending Stock	83.39	84.66
Middle Distillate Oil	124.2	123.0
Heavy Distillate Oil	<u>32.4</u>	
TOTAL OUTPUT:	333.32	322.9
<u>Overall Thermal Efficiency:</u>	61%	59%

* HHV t of coal end products.

**Blend of Illinois No. 6 and Kentucky 9/11 coals.

***Electricity is estimated at 3,413 Btu/kWh at the plant gate.

TABLE 40

HYDROGEN UTILIZATION EFFICIENCY*
FOR PHASE 0 AND MODIFIED PHASE 0 PLANTS

	<u>Phase 0 Plant</u>		<u>Modified Phase 0 Plant</u>	
	<u>lbs of H₂/</u> <u>100 lbs of Coal</u>	<u>Percent</u>	<u>lbs of H₂/</u> <u>100 lbs of Coal</u>	<u>Percent</u>
<u>Input</u>				
Coal Feed	4.96	50.2	4.96	47.5
H ₂ Consumed in the Process	<u>4.91</u>	<u>49.8</u>	<u>5.47</u>	<u>52.5</u>
Total H ₂ Input	9.87	100.0	10.43	100.0
<u>Output</u>				
C ₁ - C ₃	2.23	22.7	2.49	23.8
C ₄ - Liquid Products	4.97	50.2	5.26	50.6
Vacuum Bottoms	1.44	14.6	1.44	13.8
H ₂ , NH ₃ , H ₂ S, Fuel Gas, and Losses	<u>1.23</u>	<u>12.5</u>	<u>1.24</u>	<u>11.8</u>
Total Output	9.87	100.0	10.43	100.0

*Percent of input hydrogen in the C₄⁺ - liquid products.

23 percent of the total input for the Phase 0 plant to about 24 percent for the modified Phase 0 plant. The slight increase in hydrogen consumption is a direct result of a higher yield of C_3 fraction for the modified Phase 0 plant as compared to the Phase 0 plant. The hydrogen utilization efficiency of the modified plant, however, is similar to that of the Phase 0 plant, thus indicating a significant potential for further improvements in C_4^+ -distillate product yields.

5.4 Impacts on Economic Performance

The impacts of the minor plant modifications described in the previous subsection on the overall economic performance of the commercial H-Coal plant are assessed in this section.

The incremental capital costs associated with the required capacity expansions in the gasification and purification plant (Plant 12), the oxygen plant (Plant 15), gas plant (Plant 7), and steam plant (Plant 31) are estimated by using an 0.7 power or cost scaling factor. The incremental capital costs are summarized in Table 41. The total incremental plant costs, including contingency, is about 68 million dollars. The costs are in January, 1981 dollars, and are about two percent of the total plant costs (see Table 23, page 71).

The increased high-pressure steam and electric power consumption will proportionately increase the total gross annual operating cost for the modified Phase 0 plant. The increased steam consumption translates into approximately 151 tons per day of

TABLE 41

SUMMARY OF INCREMENTAL CAPITAL COSTS
FOR THE MODIFIED PHASE O PLANT

I. Major Process Plant Areas

<u>Plant Number</u>	<u>Area Name</u>	MM \$ <u>(January, 1981)</u>
12	Gasification and Purification	20.00
15	Oxygen Plant	4.00
7	Gas Plant	8.00
31	Steam Plant	<u>11.00</u>
Subtotal:		43.00

II. Other Cost Components

Initial Catalyst and Chemicals	1.00
Field Indirects	8.9
Engineering and Fee	<u>6.3</u>
Subtotal Incremental Plant Costs:	59.2
Contingency	<u>8.9</u>
TOTAL Incremental Plant Costs:	68.1

additional as-received, run-of-mine coal consumption in the steam plant (Plant 31). The additional coal consumption will add approximately 1.4 million dollars to the total annual operating costs. It is estimated that the modified plant will consume approximately 9,600 kWh/hr of additional electric power. The additional power consumption will add about 2.6 million dollars to the Phase 0 annual operating costs. It is assumed that the Phase 0 plant modifications will not change the labor and supplies/material components of the annual operating costs. However, local taxes and insurance component will increase in direct proportion to the total plant cost increment. A summary of incremental annual operating costs for the modified Phase 0 plant is given in Table 42. The total incremental operating costs are approximately five million dollars, which is approximately one percent of the total annual operating costs of the Phase 0 plant.

The base case expected product prices and revenues for the modified Phase 0 plant are given in Table 43. The total revenues generated from the sale of products/by-products produced in the modified Phase 0 plant in the year 1991 are about 1,860 million dollars (current dollars). The revenues generated in the same year for the Phase 0 plant were about 1,870 million dollars (see Table 31, page 87). Thus the revised C_3 -distillate product yield structure of the modified Phase 0 plant has decreased the total plant revenues only by less than one percent. This can also be illustrated by comparing the composite price (i.e., expected market value) for

TABLE 42

SUMMARY OF INCREMENTAL ANNUAL OPERATING COST
FOR THE MODIFIED PHASE O PLANT

(87% Stream Factor)

I. <u>Feedstock and Utilities Costs</u>	<u>January 1981 Dollars</u>
Coal - \$29/ton, 151 TPSD*	1,390,515
Purchased Electric Power - 9,597 kWh/hr at 3.654 ¢/kWh	<u>2,669,634</u>
Subtotal:	<u>4,060,149</u>
II. <u>Labor Costs</u>	--
III. <u>Supplies and Materials</u>	--
IV. <u>Other Costs</u>	
Local Taxes and Insurance	<u>771,900</u>
Total Incremental Annual Operating Costs:	<u>4,832,049</u>

* As received, run-of-mine coal.

TABLE 43

BASE CASE PRODUCT PRICES AND REVENUES
FOR MODIFIED PHASE O COMMERCIAL PLANT
(Current Dollars)

	Average 1991 ^(a) Rate per Calendar Day	Units	Price per Unit (Dollars)	Annual Revenue (MM Dollars)
<u>Liquid Products</u>				
Distillate Oil	18,021	Bbl	91.14	599.50
Reformate	10,854	Bbl	121.38	480.90
Propane	7,125	Bbl	74.13	192.80
Butane	6,801	Bbl	79.42	197.15
Light Gasoline	<u>2,447</u>	Bbl	87.19	<u>77.90</u>
	45,248	Bbl	93.70 ^(b) (44.8) ^(d)	1,548.25
<u>By-Products</u>				
Pipeline Gas	39,959 ^(c)	MSCF	13.76	200.70
Sulfur	461.4	Long Tons	464	78.1
Ammonia	168.2	Tons	543	<u>33.3</u>
TOTAL REVENUE				1,860.35 (891.

(a) 1991 represents the first year at full production rate.

(b) Composite price for C₃+ - 850°F liquid product.

(c) Includes about 30 percent refinery-rich gas.

(d) The numbers in parentheses are in constant 1981 dollars.

C₃+850°F liquid products produced in the Phase 0 and modified Phase 0 plants. The composite C₃+850°F liquid product price for the modified Phase 0 plant is approximately \$0.7/Bbl lower (or less than one percent lower) than the composite price for the Phase 0 plant.

The Phase 0 plant modifications will increase the total plant and operating costs slightly, and therefore, will not have any significant impact on the DCF-ROE, an indicator of the overall economic performance of the H-Coal commercial project. For example, it is estimated that the Phase 0 plant modifications will increase the total plant costs by only two percent. The sensitivity analysis, as discussed in the earlier section, has indicated that for the debt financing case, every percentage point increase in the total capital cost will decrease the DCF-ROE by about 0.3 percentage point. Thus, a two-percent increase in the capital cost will reduce the DCF-ROE only by 0.6 percent. The DCF-ROE was not sensitive to changes in annual operating costs. One percentage point increase in annual operating costs will reduce the DCF-ROE only by 0.1 percent.

6.0 CONCLUSIONS AND R&D RECOMMENDATIONS

6.1 Conclusions

The performance of the H-Coal process obtained at a smaller scale plant (PDU) is confirmed at a large-scale pilot plant operation. Therefore, the product yield structure and other process operating data collected during the selected PDU runs provide an adequate basis for projecting the technical and economic performance of the proposed H-Coal commercial plant. The differences in the product yield structure obtained during the PDU and the pilot plant runs were not significant enough to change the projected technical and economic performance of the commercial plant developed under the Breckinridge project initial effort.

The duplication of PDU operating results during the large-scale pilot plant operation reflects the successful scale-up of the H-Coal ebullating-bed reactor system. The scale-up factors used for the smaller PDU and large-scale pilot plant, along with the expected scale-up for a single commercial-scale reactor, are given in Table 44. The successful operation of the large-scale H-Coal pilot plant demonstrates scale-up factors of 7 and 73 based on reactor diameter and throughput, respectively. The corresponding scale-up factors between the pilot plant and the proposed pioneer commercial plant are 2.3 and 11, respectively. Therefore, the H-Coal commercial plant is more likely to duplicate the process performance achieved during the pilot plant operations.

TABLE 44

H-COAL REACTOR SCALE-UP FACTORS

	H-Coal Reactor Diameter	Throughput (Coal)	Scale-up Factors	
			Based on Diameter	Based on Throughput
PDU	8.5 in.	3 TPD	10.4*	240*
Pilot Plant	5 ft.	220 TPD	7.0	73.0
Commercial Plant	10-13 ft.	2,357 TPD**	2.3	11

*Represents approximate scale-up between bench-scale and PDU unit.

**Capacity for each of seven on-line reactors.

In addition to demonstrating the successful scale-up of the ebullated-bed reactor system, the pilot plant operations have also demonstrated the successful operation of critical unit operations such as the slurry mixing system, slurry heater, hydroclone system, letdown valve system, etc. The process performance and engineering data collected from the pilot plant runs will provide crucial inputs to the detailed design/engineering, construction, and startup phases of a proposed commercial plant. Thus, the technical risks associated with the first-of-kind, H-Coal commercial plant will be minimized. Technical risk is further reduced by the experience gained in scaling up the ebullated bed reactor concept to a commercial scale as well as demonstrating catalyst addition and withdrawal operation for petroleum resid hydrotreating and desulfurization.

Even though the technical risk is minimized, the economic risk associated with the proposed first of a kind project is still quite

significant. Bechtel's economic analysis, which is highlighted earlier, indicates that the proposed commercial project requires a capital expenditure of about six billion dollars. Assuming a 100-percent equity financing, the project will yield a very low discounted cash flow rate of return to attract any private investment. The debt financing, assuming a maximum of three billion dollar loan guarantee from the Synthetic Fuels Corporation, makes the proposed project economically viable. The economic viability, however, is contingent upon keeping the total capital expenditure within the above estimates. A 15-percent increase in the total capital cost will make the project economically less attractive in spite of the assumed debt leveraging.

The minor modifications required in the Phase 0 design as a result of pilot plant operations, do not alter the above conclusions regarding the economic viability of the proposed project.

6.2 R&D Recommendations

The H-Coal direct liquefaction process in its present configuration, as indicated earlier, has gone through extensive research, development, and pilot testing work. Even though the process is ready for commercial implementation, there are opportunities for selected research and development (R&D) efforts to improve the overall liquid yield and quality. These R&D opportunities are described below.

6.2.1 H-Coal Process Improvements

The proposed H-Coal commercial plant in its baseline configuration produces approximately 20,800 Bbl/day of distillate oil (400 - 850°F), which is about 47 percent of the total C₄ - 850°F liquid yield. The

bulk of this distillate oil is middle distillate with a boiling range of 400 to 650°F. The heavy oil, or vacuum gas oil (VGO) has a typical boiling range of 650 to 850°F. The properties of these distillate oil fractions are presented in the earlier section. Because of the higher nitrogen and oxygen content of these coal-based distillate oils, their compatibility with the comparable petroleum-based fractions is questionable. In addition, these oils exhibit poor stability under long-term storage conditions. Because of these constraints, HRI has conducted several feasibility studies to examine different H-Coal process configurations to maximize the production of transportation fuels.

One such configuration which seems to offer potential economic advantage over the baseline commercial plant is the extinction of vacuum gas oil fraction by recycling it to the H-Coal reactor, and thus increasing the selectivity to C₄-650°F liquid product. The C₅-650°F syncrude is subsequently upgraded to gasoline and diesel fuel by a combination of a high-severity hydrotreating and reforming unit. The extinction of VGO fraction by selective recycle is expected to be accomplished by increasing the baseline H-Coal reactor volume by about 25 percent, with a corresponding increase in reactor hydrogen partial pressure of about 15 percent.⁽¹⁵⁾ The extinction of VGO in the H-Coal reactor will require high-severity operation as opposed to the baseline syncrude operation. Because of the high-severity operation, a lesser amount of vacuum bottoms will be available from a VGO extinction mode of operation. Since the bulk of the process hydrogen requirement for the baseline plant was provided from the vacuum bottoms gasification,

additional sources of hydrogen will be required for the VGO recycle plant. It is assumed that steam reforming of C_1-C_2 gases, coupled with coal gasification, will be able to meet the conceived hydrogen deficit.

The operation of the H-Coal process in VGO extinction mode was successfully tested in bench-scale operation for selected foreign coals. However, the operability of the H-Coal process in the VGO extinction mode should be tested both at bench- and PDU-scale operations so that meaningful performance data can be obtained to confirm its potential technical and economic advantages over the baseline syncrude mode of operation.

6.2.2 General Liquefaction Studies

One aspect which calls for study is the improved management of hydrogen in converting coal to liquid and gaseous fuels. The current version consumes 23 percent of the total hydrogen in the production of C_1-C_3 hydrocarbons. A significant reduction in C_1-C_3 gas make would increase liquid fuel yields, reducing overall hydrogen consumption and possibly product cost.

Basic studies on coal liquefaction indicate that the transition from solid coal to distillate liquid fuels involves a series of sequential and concurrent reactions. Optimum reaction conditions for preferred thermal and catalytic reactions are quite diverse, which strongly suggests the need for carrying out the overall process in more than one stage.

The need to compress all the reactions into a single reactor subjected to fixed reaction conditions is an inherent limitation of the one-stage concept utilized in H-Coal and other advanced coal liquefaction processes like EDS, SRC I/II, and Kohleol.

Continued basic studies are recommended which are aimed at optimizing the various steps in the transition to clean distillates as well as establishing the optimum application of thermal versus catalytic conversion for each step, as follows:

- Coal to preasphaltenes
- Preasphaltenes to asphaltenes plus oil
- Conversion of residue to distillate
- Elimination of heteroatoms (O, N, S)
- Optimization of the addition of donor hydrogen

These studies should be coordinated with multi-stage liquefaction process developments currently under investigation by Lummus, Chevron, Kerr McGee, and HRI.

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