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LOW-SEVERITY CATALYTIC TWO-STAGE LIQUEFACTION PROCESS: ILLINOIS COAL CONCEPTUAL COMMERCIAL PLANT DESIGN AND ECONOMICS

By
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G. Wilson

September 1988

Work Performed Under Contract No. AC22-85PC80002

Hydrocarbon Research, Inc.
Lawrenceville, New Jersey

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ABSTRACT

Hydrocarbon Research, Inc. (HRI) is conducting a program for the United States Department of Energy (DOE) to evaluate a Catalytic Two-Stage Liquefaction (CTSL) Process. This program which runs through 1987, is a continuation of an earlier DOE sponsored program (1983-1985) at HRI to develop a new technology concept for CTSL. The earlier program included bench-scale testing of improved operating conditions for the CTSL Process on Illinois No. 6 bituminous coal and Wyoming sub-bituminous coal, and engineering screening studies to identify the economic incentive for CTSL over the single-stage H-Coal® Process for Illinois No. 6 coal. In the current program these engineering screening studies are extended to deep-cleaned Illinois coal and use of heavy recycle.

The results from this comparison will be used as a guide for future experiments with respect to selection of coal feedstocks and areas for further process optimization. A preliminary design for CTSL of Illinois deep-cleaned coal was developed based on demonstrated bench-scale performance in Run No. 227-47(I-27), and from HRI's design experience on the Breckinridge Project and H-Coal® Process pilot plant operations at Catlettsburg. Complete conceptual commercial plant designs were developed for a grassroots facility using HRI's Process Planning Model. Product costs were calculated and economic sensitivities analyzed.

Results of this economic evaluation show a considerable price advantage achieved by incorporating the latest CTSL Process improvements. Significant further cost reduction could be affected by increasing the efficiency in the deep-cleaning operation.

PROJECT DESCRIPTION

In September 1985 HRI initiated a two-year cost-shared program to study and develop a catalytic liquefaction process, based on a low-temperature, first-stage, catalytic two-stage process. The Statement of Work consisted of the following:

TASK 1 - PROJECT WORK PLAN

TASK 2 - EXPERIMENTAL STUDIES

Laboratory Testing

Microautoclave tests on solvents, coals, products and catalysts, and microreactor catalyst screening.

Bench-Scale Testing

Major effort on process improvement and optimization for Illinois No. 6 coal, plus evaluation of a second, alternative coal. Some of the process improvements evaluated include:

- Effects of residence time, temperature, reduced recycle solvent quantity and recycle solvent composition.
- Effect of catalyst cascading in reducing catalyst requirements.
- Effect of catalyst improvements including catalyst size, shape, promoters and pore size distribution.
- Effect of coal cleaning.
- Effect of longer residence time in the first stage.
- Effect of recycling 750°F$^+$ oils to extinction.
- Effect of a thermal second stage.

TASK 3 - ENGINEERING ANALYSIS

Data Analysis and Correlation

Economic Screening Studies

An evaluation of Wyoming coal and an updated assessment on Illinois No. 6 coal based on process improvements demonstrated in Task 2.

Conceptual Plant Designs

Plant designs on Wyoming coal and an update on Illinois No. 6 coal to provide a basis for the economic screening studies.

TASK 4 - PROJECT ADMINISTRATION AND REPORTING

In addition to normal reporting requirements four topical reports and a final report will be prepared. The topical reports are as follows:

Topical	
1	Wyoming Coal Conceptual Commercial Plant Design and Economics
2	Illinois Coal Program
3	Illinois Coal Conceptual Commercial Plant Design and Economics Update
4	Alternative Coal Studies

This report covers work performed under Task 3 on Illinois coal.

EXECUTIVE SUMMARY

INTRODUCTION

Hydrocarbon Research, Inc. (HRI) has extensive experience in the development of coal liquefaction process technology. The H-Coal® Process, which features a single catalytic ebullated-bed reactor, has been successfully developed and demonstrated at bench, PDU and large pilot plant scales. It is generally recognized, however, that single-stage coal liquefaction cannot achieve all of the multiple objectives of coal liquefaction, given the single set of operating conditions. As a result, various two-stage coal liquefaction concepts have evolved, which to varying degrees have been successful in improving coal liquefaction chemistry.

In 1981-1982, HRI conducted a series of laboratory experiments using a thermal first-stage reactor, followed by a closely-coupled, catalytic ebullated-bed, second-stage reactor. This thermal-catalytic process concept features a sequential reaction scheme of complete liquefaction in a non-catalytic system, followed by catalytic hydrogenation in a separate second stage. During this work HRI realized that there may be considerable advantages to a two-stage system that accomplishes liquefaction and hydrogenation simultaneously as opposed to sequentially. This realization formed the basis for the "New Technology" concept of Catalytic Two-Stage Liquefaction (CTSL). The concept of CTSL is to operate two closely-coupled catalytic ebullated-bed reactors in series. The first stage operates at substantially lower temperatures than the second stage.

In 1983 HRI was granted a two-year contract to investigate this process concept at laboratory and bench unit scales. The first-year of the program focused on Illinois No. 6 bituminous coal. A series of process variable studies were conducted at the bench-scale to determine the preferred operating conditions for CTSL. After the preferred operating conditions were established, a twenty-five day Demonstration Run 227-20(I-11) was carried out at constant operating conditions, to determine catalyst deactivation rates and process performance at commercial catalyst age. The demonstration run provided the basis for a comparative engineering and economic evaluation of CTSL relative to the H-Coal® Process.

The results of this study showed a 12% economic advantage for CTSL over the H-Coal® Process.

The second year of the program focused on Wyoming sub-bituminous coal. Similar to the work on Illinois No. 6 coal, a series of bench-scale process variable runs were conducted to determine the preferred CTSL operating conditions for Wyoming sub-bituminous coal. A demonstration run was then conducted at the preferred operating conditions to establish catalyst deactivation rates and process performance at a commercial catalyst age.

In 1985 HRI was granted a new two-year contract from DOE to further investigate the CTSL Process. Bench efforts have been continued aimed at improving process performance with Illinois No. 6 coal over the level demonstrated in the above mentioned studies. These efforts recently culminated in a twenty-five day Process Demonstration Run 227-47(I-27). Results of this run formulated the basis for a revised engineering estimate. This report contains the results of the engineering estimate and presents the economic analyses based on the results of the updated engineering evaluation. The Run 227-47(I-27) extinction recycle process is designated as heavy recycle operation (>750°F) and the previous process run, 227-20(I-11), as light recycle operation (>550°F).

CTSL PROCESS DESCRIPTION

Based on data developed under this program, as well as earlier work by HRI and others, productin of high yields of distillate liquid fuels of high quality from direct coal liquefaction is favored by catalytic processing in the presence of hydrogen. HRI's approachis to utilize the commercially-proven catalytic ebullated-bed reactor system in two closely-coupled stages.

A simplied flow scheme for the Catalytic Two-Stage Liquefaction Process (CTSL) is shown in Figure 1. Coal is slurried with a process-derived recycle slurry oil, mixed with hydrogen and fed to the first-stage catalytic ebullated-bed reactor. The key feature that differs from most processes is the lower first-stage temperature (<800°F) and conditions favoring early hydrogenation, directly and through the promotion of hydrogen transfer reactions in the solvent. products, characteristically are better quality than other competing direct liquefaction processes, i.e., high hydrogen/carbon ratio, higher API gravities and lower heteroatom content.

The CTSL Process conducts the conversion and hydrogenation simultaneously rather than sequentially thus capping the free radicals and reducing the formation of unstable products likely to recombine and polymerize. By operating the first stage at lower temperatures with catalyst, and high hydrogen pressure, conditions exist that favor in situ regeneration of the solvent and transfer of hydrogen to the coal and coal products as formed, reducing the chance for condensation reactions.

First stage products pass directly to the higher temperature second stage for further coal and heavy oil conversion in the presence of a high relative concentration of regenerated solvent. The optimum second stage conditions occur when coal and residuum conversion and heteroatom removal peak, without approaching a thermal severity where product dehydrogenation becomes significant.

Both reactor stages use conventional ebullated-bed hydrogenation/hydroconversion catalysts. The second-stage reactor effluent is separated into vapor and liquid-slurry streams. The vapor is cooled and further separated to recover hydrogen which is recycled back to the reactor. The reactor slurry product is depressurized and fractionated to produce high quality distillate products and the slurry is feed to solids-separation to produce a heavy 750°F$^+$ recycle slurry oil (solids-free) and a solid product (ash and unconverted coal). With the CTSL Process conditions selected for this conceptual design, near extinction of the heavy liquids recycle is achieved.

OBJECTIVES AND SCOPE OF WORK

The objective of this study is to determine the economic impact of several process improvements demonstrated in the research program since the earlier Illinois conceptual design, based on Run 227-20(I-11). These process improvements include the following:

- Reduced Coal Feed Ash Level

- Reduced Recycle To Coal Feed Ratio
- Use of Heavy Recycle Stream
- Increased Reactor Volume
- Increased distillate Liquid Yield

The Scope of Work includes:

- Development of a preliminary engineering design on the liquefaction section (Area 200) for CTSL of Illinois coal.
- Preparation of a conceptual design of a complete grassroots coal liquefaction plant using HRI's Process Planning Model.
- Calculation of products costs for CTSL of Illinois coal and overall assessment of economics including comparison with prior results on Illinois coal using light recycle.

BASIS OF DESIGN AND ASSUMPTION

Yields from liquefaction are elementally balanced based on demonstrated performance in Run 227-47(I-27) on Illinois No. 6 coal. This coal had been cleaned by heavy media washing to reduce the ash content to 5.77 W %. The CTSL distillable liquid product yield is 77.5 W % on moisture and ash-free (M.A.F.) coal, compared to 50.9 W % on M.A.F. coal for the single-stage H-Coal® Process with Illinois coal and with 64.5 W % previously demonstrated in light recycle CTSL operations.

In addition, the distillate product from the heavy-recycle CTSL operation contains a much lower proportion of heavy gas oil, with only 6.3% of the distillable product boiling above 750°F, compared to a value of 13.6% of the distillable product for the light recycle CTSL operation.

The preliminary liquefaction section design is based on HRI's experience from the Breckinridge project. Five parallel reactor trains are provided for Illinois coal. Total coal rate to liquefaction is 8,400 T/D on a dry basis. This is lower than the 9,428 T/D used for light recycle and in the Breckinridge Project design due to the higher hydrogen consumption and maximum reactor diameter considerations. However, the greater efficiency of the heavy recycle CTSL operation results in a higher yield of plant products 41,948 BPSD, compared to 34,934 BPSD for the light recycle CTSL operation.

Other design basis information is as follows:

- Oil-to-Solids Ratio in the coal feed slurry is 1.5:1. Deashing is used to provide a solids-free recycle thereby reducing total recycle liquid requirements compared to the solids containing recycle slurry used in the light recycle Illinois coal design.

- Reactor Effluent from the second stage is quenched with recycle hydrogen to reduce the potential for coking in the reactor effluent separator and in downstream equipment.

- The Slurry Mix Tank is operated at 450°F.

- Coal Feed Space Velocity is lower, 44.4 Lbs/Hr/Ft^3 catalyst (stage) compared to the prior design with Illinois coal, 67 Lbs/Hr/Ft^3 catalyst (stage).

- Catalyst Replacement Rate has been increased to 3 Lbs/Ton coal from the 1 Lb/Ton coal used in the previous light recycle study. This revised catalyst utilization is based on the apparent catalyst deactivation in the kinetic modelling of the key reactions involved applied to distribution in the reactor of catalyst of various ages in an operating system with continuous catalyst replacement.

The designs of the other areas of the conceptual commercial plant are specified consistently using HRI's Process Planning Model. The ash concentrate from deashing is gasified via partial oxidation for hydrogen production. The balance of the hydrogen required by the plant is produced via partial oxidation of the ash concentrate reject stream from deep-cleaning the coal. The whole liquid product from liquefaction is hydrotreated, and the hydrotreated heavy naphtha is catalytically reformed. The principal products from the plant are gasoline and diesel fuel.

SUMMARY OF RESULTS

Results of the conceptual commercial plant design are summarized in Table 1. Dry coal feed to the plant is 12,805 TPSD. Electric power (106 MW) is purchased. The total product rate is 41,948 BPSD. The total plant investment is $2109MM (1984$) and the product cost is $38.35/Bbl.

OVERALL ASSESSMENT OF ECONOMICS

The economics calculated are non-site specific and based on U. S. Gulf Coast investment costs and mine-mouth coal costs. This was performed to afford the consistent basis for comparison with previous studies.

With the evaluation basis selected the heavy recycle CTSL Process shows a 5% price advantage over the previous light recycle mode ($38.35/Bbl versus $40.25/Bbl). Both CTSL processes are significantly lower than the one-stage H-Coal® Process ($45.94/Bbl).

Significant further cost reduction could be achieved by increasing the efficiency of the coal cleaning operation. Thus for an efficiency increase from 70 to 80% the product cost would decrease from $38.35 to $36.69/Bbl, using steam reforming of purchased natural gas.

CONCLUSIONS AND RECOMMENDATIONS

The main conclusion from this study is that heavy recycle CTSL operation with deep-cleaned Illinois No. 6 coal presents a significant reduction in cost over light recycle operation with Illinois No. 6 coal. The product cost for the heavy recycle CTSL operation with deep-cleaned Illinois No. 6 coal is estimated to be $36.69/Bbl, compared to $40.25/Bbl for light recycle mode without coal cleaning.

SUMMARY OF RESULTS

CASE	Heavy Recycle	Light Recycle
COAL FEED, TPSD		
To Liquefaction	8,400	9,428
To Partial Oxidation	4,405	1,552
TOTAL	12,805	10,980
Purchased Electric Power, MW	106	90
LIQUID PRODUCTS, BPSD		
Gasoline	13,170	10,967
Diesel	28,778	23,967
TOTAL	41,948	34,934
BY-PRODUCTS		
LPG, BPSD	289	-
Sulfur, TPSD	329	376
Ammonia, TPSD	143	112
Ash to Disposal, TPSD	1,498	1,254
Total Plant Investment, MM$(1)	2,109	1,918
PRODUCT COST(2)		
Operating Cost, MM$/Yr	546	479
By-Product Revenue, MM$/Yr	15	12
Product Cost, MM$/Yr	531	467
$/Bbl	38.35	40.52

Notes:

(1) 1984 total erected cost at a U. S. Gulf Coast location. Includes 25% project contingency.

(2) First year product cost based on 25% equity, 15% DCF return, 10% interest on debt and 5.0% annual inflation (see Section 5).

HRI EBULLATING-BED
CATALYTIC TWO-STAGE COAL LIQUEFACTION (CTSL) PROCESS
SIMPLIFIED FLOW PLAN

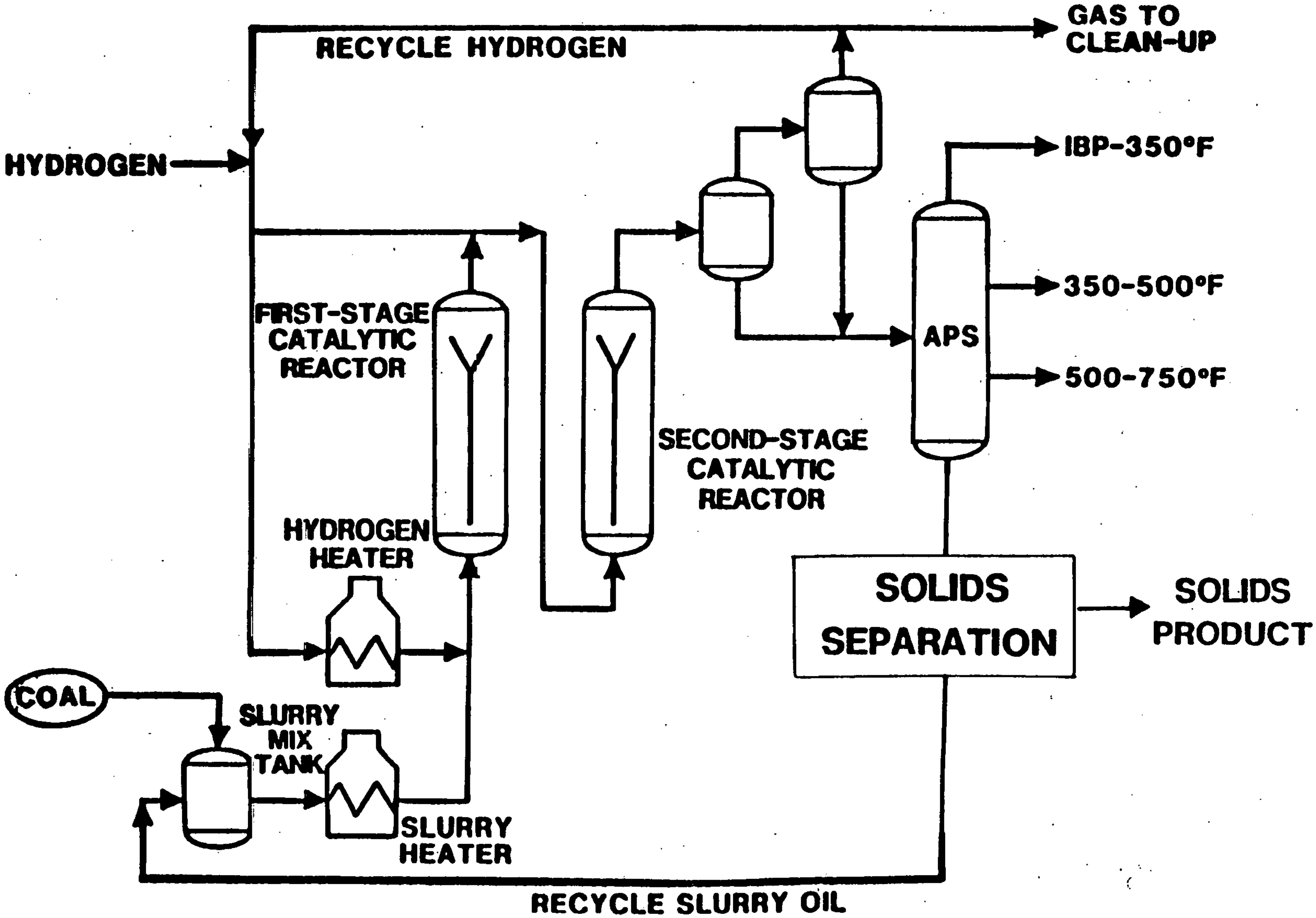


FIGURE 1

SECTION 1

OBJECTIVES AND SCOPE OF WORK

OBJECTIVES AND SCOPE OF WORK

The objective of this conceptual commercial plant design and economics study is to calculate the economics for CTSL using Illinois No. 6 bituminous coal with heavy severity. Consistent with this overall objective are the following specific objectives:

- Provide a preliminary engineering design of the CTSL plant with Illinois No. 6 coal on a comparable basis with the prior designs executed for H-Coal®, CTSL on Illinois No. 6 coal with light recycle and Wyoming sub-bituminous coal.
- Use HRI's Process Planning Model to develop an optimized conceptual commercial plant design and economics for complete a grassroots plant, including all necessary onsite and offsite facilities.
- Compare CTSL economics with prior results using Illinois No. 6 coal with light recycle.
- Identify areas for improvement in CTSL design and evaluation in the contract extension.

To meet the objectives of this study, the scope of work includes the development of a complete conceptual commercial plant design for CTSL on Illinois No. 6 coal. A preliminary engineering design of the liquefaction section (Area 200) is included in sufficient detail to support a meaningful budget-type cost estimate. A summary of work included in the preliminary engineering design is as follows:

- Design Basis
- Battery limits include coal slurry preparation, preheat, reaction, primary separation, hydrogen recycle, recycle slurry preparation and product fractionation.
- Process Description
- Overall Material Balance

- Utilities and Catalyst Consumption
- Major Equipment Specifications
- Battery limits estimates of major equipment costs.

HRI's Process Planning Model is used to optimize the remaining areas of the plant including: coal preparation, hydrogen manufacture, emission and effluent control systems, product upgrading, utilities and general offsites. Optimized economics are determined by the computer model simulations and the overall product costs are calculated.

SECTION 2

BASIS OF DESIGN AND ASSUMPTIONS

BASIS OF DESIGN AND ASSUMPTIONS

The CTSL conceptual commercial plant design is based on performance demonstrated in bench-scale Run No. 227-47(I-27)(1) on Illinois No. 6 coal (Burning Star Mine). This twenty-five-day demonstration run was conducted at constant operating conditions, progressive with temperature adjustment to maintain minimum heavy oil yields, determined to be an improved set of operating conditions from prior bench scale screening. The design coal feed analysis is shown in Table 2. Yields for CTSL are based on Period 21 from Run 227-47(I-27). CTSL Process yields are also shown based on performance previously demonstrated in Run 227-20(I-11)(2). CTSL Process yields and process performance are compared in Table 3.

CTSL distillable liquid product yield (C_4-975°F) is 77.5 W % on M.A.F. coal compared to 64.5 W % on M.A.F coal previously demonstrated. This represents an increase in liquid product yield in excess of 20%. In addition, the distillate product from the heavy recycle CTSL operation contains a much lower proportion of heavy gas oil, with only 6.3% of the distillable product boiling above 750°F, compared to a value of 13.6% of the distillable product for the light recycle CTSL operation. The distillate sulfur and nitrogen levels are reduced. Those changes are reflected in lower downstream treating costs for the coal liquefaction products. Overall hydrogen consumption is higher than previously demonstrated and the hydrogen efficiency (weight ratio of C_4-975°F liquid yield to hydrogen consumption) is slightly less (9.9 versus 10.7) than previously demonstrated.

The preliminary design of the liquefaction sections is based on HRI's experience from the Breckinridge Project(3). The Breckinridge Project design, completed in 1982, was a site-specific design of a pioneer (first) commercial-scale coal liquefaction plant using the H-Coal® Process on Illinois No. 6 coal. HRI was responsible for the design of the liquefaction section of the plant. The design included eight parallel reaction trains. Subsequent studies by Bechtel showed a plant with four parallel reaction trains to be preferred as a minimum economic plant size. This study is based on providing a slightly lower coal feedrate to liquefaction than in the four train light recycle Illinois No. 6 coal design. The capacity of the reactor trains is set by providing maximum-sized (diameter) reactors that can be shop-fabricated using conventional techniques for heavy-walled vessels. Other design basis information is as follows:

- <u>Oil-to-Solids Ratio</u>
  The oil-to-solids ratio in the coal feed slurry is 1.5:1. Deashing is included in the liquefaction section so that a solids-free recycle slurry oil is provided.

- <u>Catalyst Replacement Rate</u>
  The catalyst replacement rates were rigorously calculated for both the first and second stages based on process kinetic models derived from bench-scale data. (See Addendum)

- <u>Quench of Reactor Effluent</u>
  The reactor effluent from CTSL (second-stage) is quenched with recycle hydrogen. This quench was not included in the original Breckinridge Project design. However, it was thought to be necessary to prevent coking in the reactor effluent separator at the Catlettsburg H-Coal® Pilot Plant(4).

Design of the other areas of the conceptual commercial coal liquefaction plant are specified using HRI's Process Planning Model(5). HRI's Process Planning Model is based on the latest process and economic information available for all onsite and off-site areas required in a complete grassroots coal liquefaction facility. Those areas include:

Area	Description
100	Coal Preparation
200	Liquefaction
300	Partial Oxidation
400	Oxygen Plant
500	Product Treating
600	Product Upgrading
700	Utilities
800	Tankage
900	General Offsites

These areas are described in detail in later sections of this report. Pertinent design basis information for the major onsite process areas is summarized as follows:

- Hydrogen Production - The ash concentrate stream from deashing in the liquefaction section is fed to partial oxidation as a water slurry for hydrogen production. Also supplied to partial oxidation is the ash concentrate reject stream from deep cleaning the coal. The balance of the hydrogen is produced via steam reforming a portion of the product gas.

- Product Upgrading - The whole liquid product from liquefaction is hydrotreated and the hydrotreated heavy naphtha (180-350°F) is catalytically reformed. The prime products from the plant include unleaded gasoline and No. 2 diesel fuel.

TABLE 2

DESIGN COAL FEED ANALYSIS[1]

Coal Type: Illinois No. 6 (Burning Star Mine)
Deep Cleaned By Heavy Media Washing

PROXIMATE ANALYSIS, W % DRY COAL	
Volatile Matter	40.30
Fixed Carbon	53.93
Ash	5.77
TOTAL	100.00

ULTIMATE ANALYSIS, W % DRY COAL	
Carbon	73.92
Hydrogen	4.93
Nitrogen	1.51
Sulfur	2.75
Ash	5.77
Oxygen (by difference)	11.12
TOTAL	100.00

HEATING VALUE (HHV), BTU/LB DRY	13,181
MOISTURE, W % ON COAL FEED TO LIQUEFACTION	3.08

SULFUR FORMS, W % DRY COAL	
Organic	2.02
Pyritic	0.72
Sulfate	0.01
TOTAL	2.75

(1)From Run 227-47(I-27)

OVERALL YIELDS AND PROCESS PERFORMANCE

W % DRY COAL	HEAVY RECYCLE(1)	LIGHT RECYCLE(2)
H_2S	2.50	2.43
NH_3	1.38	0.98
H_2O	10.57	10.52
CO_x	0.10	0.42
C_1-C_3	8.44	5.87
C_4-400°F	21.23	17.72
400-650°F	30.14	24.26
650-750°F	17.02	7.57
750-975°F	4.59	7.80
975°F$^+$	2.89	11.86
Toluene Soluble	1.82	8.88
Toluene Insoluble	1.07	2.98
Unconverted Coal	2.77	4.88
Ash	5.77	11.05
TOTAL	107.40	105.36
PROCESS PERFORMANCE		
C_4-975°F		
W % M.A.F. Coal	77.5	64.5
B/T M.A.F. Coal	5.0	4.1
C_4-750°F, W % M.A.F. Coal	72.6	55.7
Coal Conversion, W % M.A.F.	97.1	94.5
975°F Conversion, W % M.A.F.	94.0	81.2
Hydrogen Efficiency	9.9	10.7
PRINCIPAL OPERATING CONDITIONS		
Coal Feed,		
Lbs/Hr/Ft^3 Catalyst (Stage)	44.4	66.7
Solvent Recycle, Lbs/Lb Coal	1.15	2.82
Residual Oil In Solvent, W %	47.1	30.4
Temperature, °F		
First Stage	767	750
Second Stage	823	827
Hydrogen Partial Pressure, Psi	1900	1900

(1)Run 227-47(I-27)
(2)Run 227-20(I-11)

SECTION 3

PLANT CONFIGURATIONS AND OVERALL DESIGNS

PLANT CONFIGURATIONS AND OVERALL DESIGNS

Complete, grassroots conceptual commercial plant designs are developed using HRI's Process Planning Model. Plant facilities which are specified include:

Area	Description
100	Coal Preparation
200	Liquefaction
300	Partial Oxidation
400	Oxygen Plant
500	Product Treating
600	Product Upgrading
700	Utilities
800	Tankage
900	General Offsites

A detailed description of each of these plant sections is provided in Section 4 - Process and System Descriptions of this report.

Figure 2 presents a block flow diagram of the major processing areas. Coal is received and prepared in Area 100 - Coal Preparation. The deep-cleaned coal is sent to Area 200 - Liquefaction. Ash concentrate from deashing is fed to Area 300 = Partial Oxidation for production of hydrogen. The ash concentrate reject stream from deep cleaning the coal in Area 100 is likewise sent to Partial Oxidation - Area 300. Oxygen required for partial oxidation is produced in Area 400 - Oxygen Plant. The hydrogen produced goes to liquefaction and Area 600 - Product Upgrading. The liquid product from liquefaction is further processed in Area 600 - Product Upgrading. In product upgrading the total distillate liquid product is hydrotreated and the hydrotreated heavy naphtha is catalytically reformed. Purge gases and sour water from liquefaction and product upgrading are treated in Area 500 - Product Treating, where C_4s are recovered for blending to gasoline, as well as sulfur and ammonia by-products. The prime products from the plant are a high octane unleaded gasoline and a No. 2 diesel fuel.

OVERALL MATERIAL BALANCE

The overall plant material balance is summarized in Table 4. The dry coal feed to liquefaction is 8,400 TPSD. 13,170 BPSD of gasoline and 28,778 BPSD of No. 2 diesel fuel are produced, for a total liquid product rate of 41,948 BPSD or 4.99 Bbl/Ton of dry coal to liquefaction.

The by-products recovered include LPG, sulfur, ammonia and ash. Sulfur, LPG and ammonia provide revenue credits, while a cost is incurred for ash disposal.

Liquid product qualities are presented for the gasoline and diesel products in Table 5. The gasoline produced is high octane (90 (R+M)/2) unleaded. Lower octane ratings could be obtained by reducing the catalytic reformer severity (set at 103 RON). Coal derived naphthas, however, possess excellent reformability, consequently little savings are realized by reducing reforming severity. This gasoline would probably be sold to a refiner, for blending with lower octane petroleum derived gasoline. The diesel fuel has a cetane number greater than 40 with essentially no sulfur or nitrogen.

HYDROGEN BALANCE

The overall plant hydrogen balance is shown in Table 6. Presented are hydrogen consumption in liquefaction and upgrading, as well as purge and solution losses, and hydrogen production in partial oxidation of ash concentrate and steam reforming of light gases. Hydrogen consumption per barrel of liquid product is 6,580 SCF/Bbl. Overall 83% of the hydrogen is consumed in liquefaction, 14% in upgrading and 3% represent physical losses to fuel gas.

Eight percent of hydrogen is produced via partial oxidation of the ash concentrate from liquefaction, 90% via partial oxidation of the ash-concentrate reject stream from deep-cleaning coal and the balance by steam reforming of light gases.

THERMAL EFFICIENCY

Thermal efficiency is defined as the percentage of the plant energy inputs (coal and electric power) which leave the plant as products and by-products. The thermal efficiency is calculated in Table 7. The inputs to the plant are the coal feed to liquefaction and to partial oxidation, as well as purchased electric power (362 GBtu/SD).

Outputs from the plant include gasoline, diesel, LPG, sulfur and ammonia. The outputs total 246 GBtu/SD. As a result, the thermal efficiency is 68.0%.

UTILITIES SUMMARY

A summary of the utility usages is provided in Table 8. All utilities are internally supplied with the exception of purchased electric power.

CAPACITY OF ONSITE PROCESS UNITS AND OFFSITE UNITS

The capacities of the onsite process units and offsite units are summarized in Tables 9 and 10.

BLOCK FLOW DIAGRAM OF MAJOR PROCESSING AREAS

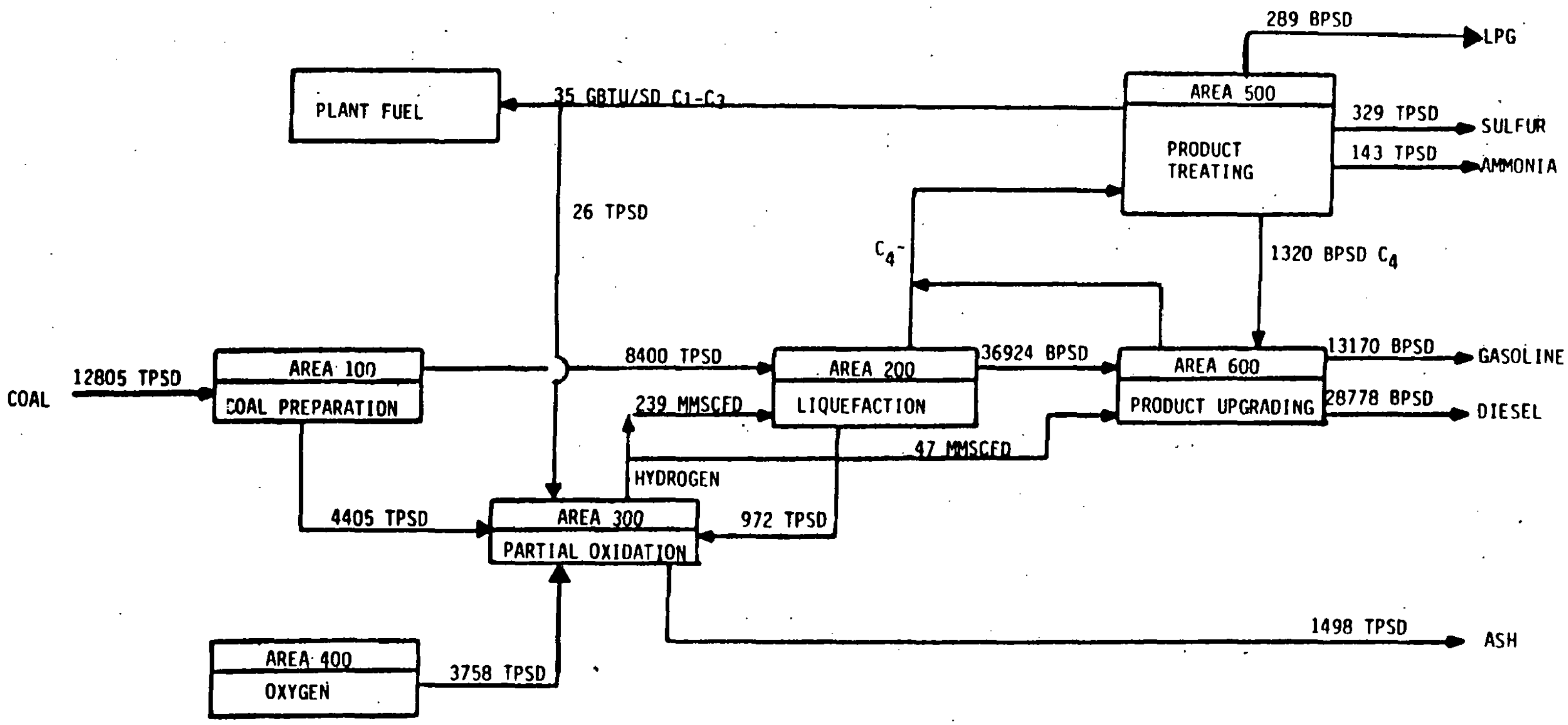


FIGURE 2

OVERALL MATERIAL BALANCE

	HEAVY RECYCLE	LIGHT RECYCLE
COAL FEED, TPSD		
To Liquefaction	8,400	9,428
To Partial Oxidation	4,405	1,552
TOTAL	12,805	10,980
Purchased Electric Power, MW	106	90
LIQUID PRODUCTS, BPSD		
Gasoline	13,170	10,967
Diesel	28,778	23,967
TOTAL	41,948	34,934
B/T Dry Coal to Liquefaction	4.99	3.71
B/T Dry Coal Total	3.28	3.18
BY-PRODUCTS		
LPG, BPSD	289	-
Sulfur, TPSD	329	376
Ammonia, TPSD	143	112
Ash to Disposal, TPSD	1,498	1,254

TABLE 5

PRODUCT QUALITIES

GASOLINE	
°API	48.7
RON (Clear)	95
RVP, psia	11.5
R+M/2, (Estimated)	90
DIESEL	
°API	34.7
Cetane Number	> 40

TABLE 6

HYDROGEN BALANCE

HYDROGEN CONSUMPTION, MMSCFD	
Liquefaction	236
Upgrading (Net)*	40
Purge and Solubility Losses	10
TOTAL	286
Hydrogen Consumption, SCF/B of Liquid Products	6580
HYDROGEN PRODUCTION, MMSCFD	
Partial Oxidation	
Bottoms	24
Coal	258
Light Gases (C_1-C_3)	4
TOTAL	286

*Net hydrogen consumption. Hydrogen consumed in hydrotreating less hydrogen produced in catalytic reforming.

TABLE 7

THERMAL EFFICIENCY

INPUTS, GBTU/SD(1)	
Coal	
To Liquefaction	221
Supplemental	116
Electric Power(2)	25
TOTAL	362
OUTPUTS, GBTU/SD(1)	
Gasoline	72
Diesel	168
LPG	1
Sulfur	3
Ammonia	2
TOTAL	246
THERMAL EFFICIENCY, %	68.0

(1)G = 10^9 (billions).

(2)Coal Equivalent. Assumes coal fired boiler at 10,000 Btu of coal per Kw of electric power.

TABLE 8

UTILITIES SUMMARY

ELECTRIC POWER, MW	
Purchased	106
Via Steam Turbine Drivers	66
TOTAL	172
600 Psig Steam, MLb/Hr	93
Cooling Water, MGPM	73
Process Fuel, GBtu/SD	35
Raw Water, MGPM	3

TABLE 9

CAPACITY OF ONSITE PROCESS UNITS

AREA	CAPACITY UNITS	
200 - Liquefaction	TPSD Dry Coal	8,400
300 - Partial Oxidation	MMSCFD Hydrogen	286
400 - Oxygen Plant	TPSD Oxygen	3,758
500 - Product Treating		
Acid Gas Removal	TPSD H_2S plus CO_2	34
Sour H_2O Stripping	TPSD Sour H_2O	4,805
Sulfur	TPSD Sulfur	329
Light Ends	TPSD C_4	26
600 - Product Upgrading		
Hydrotreating	BPSD Raw Coal Liquid	37,623
Catalytic Reforming	BPSD 180-350°F	8,817

TABLE 10

CAPACITY OF OFFSITE UNITS

AREA	CAPACITY UNITS	
100 - Coal Preparation	TPSD Dry Coal	12,805
800 - Utilities		
Power Distribution	Mw	106
Steam Generation	MLb/hr	93
Cooling Water	MGPM	73
900 - Tankage		
Liquid Products	BPSD	41,948
Waste Solids Handling	TPSD	1,498
LPG	BPSD	289
1000 - General Offsites	TPSD Dry Coal	12,805

SECTION 4

PROCESS AND SYSTEMS DESCRIPTIONS

PROCESS AND SYSTEMS DESCRIPTIONS

This section provides a description of the processes and systems contained in each of the functional areas. The functional areas in the conceptual commercial plant designs are as follows:

AREA		
100	-	Coal Preparation
200	-	Liquefaction
300	-	Partial Oxidation
400	-	Oxygen Plant
500	-	Product Treating
600	-	Product Upgrading
700	-	Utilities
800	-	Tankage
900	-	General Offsites

Area 200 is described in considerable detail, including major equipment specifications, based on the preliminary design developed for this study. Other sections of the plant are duty specified and overall designs are developed based on HRI's Process Planning Model. HRI's Process Planning Model uses state-of-the-art computer modelling techniques to develop optimized conceptual commercial plant designs and economics for a complete, grass-roots coal liquefaction facility. The process information contained in the model is based on the most recent commercial liquefaction plant designs and also reflects the learnings from the large pilot plant projects completed.

AREA 100 - COAL PREPARATION

Coal (Illinois No. 6) is received at the plant via rail car. Coal is conveyed to live storage piles for three days normal use. A thirty-day emergency coal storage pile (dead storage) is maintained. Mobile equipment is used to reclaim coal from dead storage to the active storage pile.

Coal is reclaimed from live storage and conveyed to coal storage surge bins ahead of the mills. Coal is simultaneously crushed and dried in roller mills. The hot combustion gases are filtered in a bag house and discharged to the atmosphere. Dry coal is delivered to coal storage bins at a nominal minus 30 mesh particle size.

Facilities in coal preparation included equipment required for deep cleaning the coal by heavy media washing. The ash content of coal fed to liquefaction is thereby reduced from 11.7 to 5.77 W %. At the same time an ash concentrate reject stream, containing 23 W % ash, is sent to Partial Oxidation - Area 300.

Material balance and utility summaries are provided in Tables 11 and 12, respectively.

TABLE 11

AREA 100 - COAL PREPARATION

Material Balance Summary

	POUNDS PER HOUR		
		PRODUCTS	
	Feed(1)	To Liquefaction(2)	To Partial Oxidation
M.A.F. Coal	942,300	659,610	282,690
Ash	124,810	40,390	84,420
TOTAL	1,067,110	700,000	367,110

(1)Dry Coal Basis. As received coal contains approximately 10 W % moisture.

(2)Dry Coal Basis. Crushed and dried coal contains approximately 3 W % moisture.

TABLE 12

AREA 100 - COAL PREPARATION

Utilities Summary

Electric Power, Kw	7,980
Cooling Water, Gpm	132
Process Fuel, MMBtu/Hr	266

AREA 200 - LIQUEFACTION

PROCESS DESCRIPTION

CTSL converts coal into liquid hydrocarbon products by utilizing two catalytic ebullated-bed reaction stages in series. Both reaction stages operate at lower temperatures than the conventional single stage H-Coal® process, with the first of the two stages at a lower temperature than the second.

Figure 3 presents a simplified block flow diagram of the liquefaction area. Coal is mixed with recycled oil and the combined slurry is pumped to reaction pressure, preheated and then combined with make-up and recycled hydrogen. Reaction is carried out in the two-stage catalytic ebullated-bed reactors. Reactor effluent is separated into a hydrogen-rich vapor stream which is returned to the reactor, a hydrocarbon liquid stream which is sent to fractionation and a slurry stream which is sent to solids separation. The raw products leaving liquefaction, namely naphtha, light and heavy distillates, are prepared in the fractionator which also forms the fractionator bottoms recycle oil for mixing the coal. Finally, the reactor slurry is separated into a solids-free recycle oil and a reject solids-rich stream which is sent to partial oxidation.

This CTSL plant is designed to convert 8,400 Tons/Day of Illinois No. 6 coal. The reaction area consists of five identical trains each handling 1,680 Tons/Day. The reactor train capacity is set to correspond to the maximum diameter vessel, Breckinridge and the original Illinois coal CTSL designs.

The reactor trains are identified by series 100, 200, 300, 400 and 500 equipment numbers. Reactor effluent is first processed in the separation equipment and then fed to downstream units.

A catalyst handling system is provided for each reactor train. Thus, there are five catalyst handling trains: Section 150, 250, 350, 450, and 550.

Slurry streams from the separation equipment of the five reactor trains are combined and processed in a single Recycle Oil Preparation train, Section 700. This section prepares a solids-free recycle stream containing a sufficient quantity of residuum to effect the desired level of conversion. The net yield of residuum and solids is separated in this section and fed to the gasifiers in Area 300.

One Distillate Separation train, Section 600, is provided to process the distillate streams from the five reactor trains and stripper overhead vapor from Section 700. This Distillate Separation train recovers valuable distillate products and prepares an oil stream to be recycled for slurrying fresh coal.

Material balance and utility summaries are shown in Tables 13 and 14. Major equipment specifications are shown in Tables 15 through 30.

SECTIONS 100, 200, 300, 400, 500 - REACTION

Sections 100, 200, 300, 400, and 500 are comprised of five identical trains of reaction equipment. Each section processes 1,680 Tons/Day of Illinois No. 6 coal for a combined plant capacity of 8,400 Tons/Day. Each of the five identical trains employs the same processing schemes which consist of the following steps:

1. Coal Slurry Preparation
2. Coal Slurry Pumping
3. Hydrogen Compression
4. Preheating
5. Reaction
6. Separation

These steps are described below for a typical train of the five parallel reaction trains. Reference is made to PFD No. E-12-1-01A for these sections (Exhibit 1).

COAL SLURRY PREPARATION

The coal slurry preparation step prepares the slurry feed to the reactors. The equipment required in this step include a slurry mix tank, an eductor system and its auxiliaries, and several heat exchangers.

Slurry Mix Tank

The slurry mix tank provides a means of mixing finely divided coal with oil to obtain a pumpable slurry mixture. The drum also provides surge capacity for the reactor feed.

Fresh, ground, and dried coal is delivered to the slurry mix tank (Q-101 to Q-501) through the fresh coal feed chutes. The fresh coal, available at approximately 170°F, is slurried with two process derived streams. The first stream originates from the solid-liquid separation unit, S-701 in Section 700. This solids-free stream contains the residuum recycle material required to obtain the desired level of conversion. The second slurrying stream, the fractionator bottoms obtained in the Fractionation section, is recycled in order to meet the required recycle oil composition and the oil to solids ratio necessary to form a pumpable slurry mixture. Part of the fractionation bottoms stream is used as a make-up oil to the eductor motive liquid. The remainder is delivered to the slurry mix tank.

The slurry mix tank is designed to operate at a temperature of approximately 450°F. From a thermal efficiency point of view, it is advantageous to operate at temperatures higher than 450°F. However, laboratory data indicate that above approximately 450°F, the recycle residuum material tends to increase in viscosity and precipitate heavy components. In order to maintain the mix tank at this temperature the liquid purge stream from the bottom of the eductor separator (Q-102) is heated by heat exchange with eductor make-up liquid in M-110 before being recycled back to slurry mix tank.

It is expected that some residual moisture in the fresh coal will be vaporized. These vapors are vented and condensed in the eductor system. The condensed hydrocarbon vapors are returned to the slurry mix tank and the condensed steam is used as wash water for the desalter in Section 600.

Eductor System

Contacting the relatively cold coal which contains residual moisture with hot recycle slurry and oil material results in the formation of steam and hydrocarbon vapors. These vapors, if not properly vented and condensed, would result in a build-up of back-pressure in the slurry mix tank. Vapors would be forced through the coal feed chutes and would then condense in the coal feed areas where the condensate would cause coal to agglomerate and plug up the feed lines.

To alleviate these conditions, the CTSL Process utilizes an eductor-scrubber system that has proven effective at the H-Coal® Pilot Plant at Catlettsburg. The system includes the eductor (S-101 to S-501) an eductor separator (Q-102 to Q-502) and an eductor circulating pump (J-102 to J-502).

The eductor, using a recirculating stream of hydrocarbon liquid similar in composition to the recycle oil, serves the following functions simultaneously:

- Entrains and compresses the slurry mix tank vapor vent
- Cools the slurry mix tank stream thereby condensing a significant portion of the vaporized hydrocarbons.
- Washes out any coal dust particles entrained in the vent stream.

The eductor motive liquid is supplied under flow control, from the eductor circulating pump. The motive liquid entrains Q-101 vent stream and the mixture discharges to the eductor separator (Q-102). In Q-102, condensed hydrocarbons and entrained coal dust are separated from the vapor stream. The vapor stream, which is essentially steam, is condensed in the upper packed zone of Q-102.

In order to prevent accumulation of condensed hydrocarbons and coal dust particles in the motive liquid, a slipstream of the eductor separator liquid is purged back to the slurry mix tank. Correspondingly, a fresh make-up stream of recycle oil is bled into the eductor motive liquid to maintain inventory balance. This eductor make-up oil stream is cooled by generating 200 psig steam in M-109 and heat exchange with eductor bottom purge stream in M-110. Before being bled into the eductor motive liquid, this eductor make-up oil stream is further air-cooled. The temperature of the eductor make-up oil is set such that no additional cooling is required to achieve a reasonable eductor separator bottoms temperature. Low temperature at the bottom of the eductor separator, while reducing vent losses, would increase the risk of water condensing in this section. Since the density of water is almost identical to that of the motive liquid, separation by density difference would not be feasible.

The upper section of the eductor separator consists of a packed pumparound zone designed to condense steam and light hydrocarbon vapors. The net material is withdrawn, under level control, to Q-103, where condensed sour water is separated from hydrocarbons, then used as wash water for Section 600 - Desalter.

COAL SLURRY PUMPING

The reactor slurry feed from the slurry mix tank is pumped to reactor pressure by a booster and high pressure pump system. Booster (centrifugal) pumps (J-101A/B/C) provide positive suction pressure to the high pressure (reciprocating) slurry feed pumps (J-105A/B/C). Two booster and two high pressure pumps, each handling 50% of the flow, are used to pump the slurry to the reactor pressure via slurry preheaters. A third booster and a third high pressure pump are used as spares.

Using two independent pumps at all times insures slurry flow to the reactor, even if one pump fails. Availability of slurry flow to the reactors is of major importance in the operation of the CTSL process. This is because the slurry feed absorbs the heat of reaction released in the reactors, and thus controls the reactor temperature. Loss of reactor feed would lead to vaporization of reactor liquid, eventually causing the ebullating pumps to lose suction. Loss of the ebullating pumps will cause the catalyst bed to slump and reactor operation then resembles that of a fixed-bed reactor.

Each of the slurry booster pumps (J-101) is sized to provide about 50% of the net forward stream as recycle to the slurry mix tank. This recycle stream ensures constant movement in the slurry transfer lines, thus avoiding the possibility of solids settling and lines plugging.

HYDROGEN COMPRESSION

Make-up hydrogen from the Gasification section and recovered hydrogen from the Hydrogen Purification section are combined in the make-up hydrogen suction knockout drums (Q-116). This combined hydrogen stream is compressed to reactor pressure by six two-stage reciprocating compressors (R-101-1 A/B to R-101-6 A/B) operating in parallel. Four of these compressors are used during normal operation and two are provided as spares. Each operating compressor is capable of handling 25% of the combined hydrogen feed flow to the five reactor trains. Thus, one make-up hydrogen compression system, consisting of four operating and two spare two-stage compressors, services the five parallel reactor trains. Each compressor discharges to a common manifold from which hydrogen feeds each train under flow control.

The high pressure hydrogen stream to each reactor train is combined with a recycle hydrogen stream from the recycle hydrogen compressor (R-102 to R-502) servicing that train. The total hydrogen stream then feeds the reactors via a heat exchange and preheating circuit as described below.

PREHEATING

In the CTSL Process, with two reaction stages, two separate heaters are provided to heat the reactor feed streams. One is a vapor phase heater to preheat most of the feed hydrogen to the first stage. The other is a slurry heater designed to preheat a mixture of the coal slurry feed and a portion of the feed hydrogen.

Hydrogen feed gas is first preheated by exchange against the reactor effluent vapor in M-105A/B. A portion of this hydrogen feed is injected into the coal slurry feed to the slurry heater in order to prevent coking. The slurry mixture is then heated in the slurry heater (L-101 to L-501). The first-stage reactor temperature is controlled by adjusting the slurry heater outlet temperature. The outlet temperature of the slurry heater is controlled by adjusting the rate of fuel fired to the heater.

The remaining hydrogen stream is heated to reactor temperature in the hydrogen heater (L-102 to L-502). As in the slurry heater, the outlet temperature of this heater is controlled by adjusting the rate of fuel fired. Enough extra capacity is designed into these two feed heaters to assure flexibility in controlling the operating temperatures.

REACTION

In the CTSL Process, coal conversion is achieved in two catalytic ebullated-bed reaction stages in series. The first-stage reactor operating temperature is lower than the second-stage reactor temperature. However, both reaction stages operate at a lower temperature than the conventional H-Coal® single stage reaction temperature. Using two stages rather than a single stage to achieve conversion increases the overall yields of distillate material. The first stage promotes hydrogenation, while the second stage enhances the kinetics of hydroconversion. The net result is an increase in the overall yields and an improvement in hydrogen selectivity to liquid distillate products.

The CTSL reactors (K-101 to K-501 and K-102 to K-502) incorporate the principle of ebullating bed operation as successfully demonstrated in the Catlettsburg H-Coal® pilot plant and in several commercial H-Oil® units currently operating throughout the world.

Coal-slurry-hydrogen mixture from the slurry feed heater along with preheated recycle hydrogen are introduced at the bottom of the first-stage CTSL reactor. The entire mass inside of the reactor is held in a fluidized ebullated state by recirculating liquid from the top of the reactor through the recirculating ebullating pump (J-106) and back into the bottom of the reactor. By keeping the reactor contents in an ebullated state, there is very little temperature difference across the reactor, and the exothermic heat liberated by the hydrogenation reactions is removed as sensible heat rise of the net coal-slurry feed stream. Slurry feed inlet temperature is controlled at a temperature sufficiently below reactor temperature to maintain the reactor in heat balance. In addition to the coal particles, nickel-molybdenum catalyst pellets are also held in suspension in the first-stage CTSL reactor. Since the coal particles are much finer than the catalyst pellets, a separation can be made between these solids such that the coal and ash particles are removed with the liquid-gaseous reactor effluent products, while the catalyst remains behind in suspension in the reactor. Control of the expansion of the ebullated catalyst bed is obtained by varying the circulation rate through the ebullating pump with the gas velocity held constant.

Effluent from the first-stage CTSL reactor is quenched with cold hydrogen from the recycle hydrogen compressor, and introduced into the bottom of the second-stage CTSL reactor. The interstage quench hydrogen flowrate is controlled to maintain the reactor in heat balance. Second-stage reactor operation and bed ebullation is essentially the same as that of the first-stage reactor.

The CTSL reactors are lined internally with 6" castable refractory as protection from temperature excursions which could possibly occur under abnormal upset conditions.

SEPARATION

The reactor effluent is separated into its vapor and liquid phases outside the reactor. Separation is achieved at three consecutive pressure levels.

- The high pressure level operates at essentially the reactor pressure. The hydrogen-rich stream, after the cooling, separation and water wash steps, is recycled back to the reactor.

- The intermediate pressure level operates at the same pressure level as the Hydrogen Purification Unit (HPU). Recovered hydrogen and other light gases from the intermediate pressure flashes can be fed directly to the HPU without further compression.

- A low pressure level which removes dissolved light gases plus hydrogen which would otherwise overload the downstream fractionation system. After removal of H_2S from the flash gas, the low pressure gas is blended into the fuel gas system.

High Pressure Level Flash Section

The effluent from each second-stage CTSL reactor is separated into its vapor and slurry phases outside the reactor in a separate reactor effluent separator vessel (Q-104). While in the vessel, the slurry is quenched by direct contact with hydrogen from the recycle gas compressor. Hydrogen fed through a sparger achieves equilibrium with the reactor effluent slurry. This quenching method has proven effective in preventing coke formation in the reactor separator vessel at the H-Coal® Pilot Plant at Catlettsburg. The slurry is then let down in pressure and fed directly to the intermediate pressure slurry drum (Q-107).

Vapor from Q-104 is cooled by exchange against hydrogen in exchanger M-105A/B as previously described. M-105A/B has two shells in series.

Condensate produced in M-105 is then separated from the hydrogen rich vapor stream in the high pressure hot condensing drum (Q-105). This intermediate flash drum operates at a relatively high temperature so as to 1) prevent chloride salts from precipitation and 2) prevent steam from condensing out in the presence of heavy hydrocarbons. Heavy hydrocarbons are removed prior to injecting wash water in a downstream exchanger as these heavy hydrocarbons could cause emulsion problems that could create difficulties in the downstream oil-water phase separation. The heavy hydrocarbons, collected in Q-105, are flashed adiabatically to the intermediate pressure level.

Vapor from Q-105 is water-washed to dissolve ammonium chloride, ammonium sulfide and bicarbonate salts which might form as the reactor vapor is cooled. The vapor-wash water mixture is then air-cooled in the reactor effluent cooler (M-106). The mixture leaving M-106 is separated into three phases (vapor, hydrocarbon liquid and water) in the high pressure cold condensing drum (Q-106).

Vapor from Q-106 is compressed to the reactor pressure via the recycle hydrogen compressor (R-102). The compressed hydrogen is then split into recycle, reactor interstage quench and reactor effluent quench gas streams. The reactor interstage quench hydrogen is used to quench the first-stage reactor effluent as previously described. The reactor effluent quench hydrogen is used to cool the second-stage reactor effluent slurry. The recycle hydrogen is sent back to the reactor after being preheated as previously described. A portion of the vapor leaving Q-106 is taken as a vent to prevent build-up of methane, nitrogen and other non-condensibles in the recycle hydrogen stream. This vent stream is let down in pressure and fed to a unit for hydrogen recovery. The use of a hydrogen recovery unit minimizes fresh make-up hdyrogen requirements.

The three phase mixture entering Q-106 is separated in the settling zone of the drum into a heavy water phase, a light oil phase which floats on top of the water, and a vapor phase which exists above the liquid phases. The oil phase overflows a vertical baffle and is accumulated in the oil compartment of the drum. Water is removed by means of a boot in the phase separating zone of the drum. An interface level controller controls the withdrawal rate of the water. The sour water is pressured to accumulator Q-114 where it is depressured to essentially atmospheric pressure, mixed with water wash streams from the intermediate and low pressure flash sections, and pumped to sour water stripping.

Hydrocarbon liquid from the oil side of Q-106 is depressured to the oil side of Q-109, which is a similar phase separating drum operating at the intermediate pressure level. Oil from Q-109 is subsequently depressured to the oil side of the low pressure cold condensing drum (Q-112) and then pressured to Section 600 for separation and recovery of hydrocarbons.

Intermediate Pressure Level Flash Section

Feed to the intermediate pressure flash section consists of the depressurized reactor effluent slurry from Q-104, plus the heavy and light hydrocarbon condensate streams let down from the high pressure flash section. The primary purpose of this intermediate pressure flash section is to recover a portion of the hydrogen dissolved in the reactor effluent streams at a pressure sufficiently high to be fed directly to a hydrogen recovery unit without further compression.

The reactor effluent slurry is flashed adiabatically in the intermediate pressure slurry drum (Q-107). Because of the solids' erosive characteristics, dual valves are provided for depressuring reactor effluent slurry. The valves are installed to enable one valve to be taken out of service while the other valve is in service.

Vapor from Q-107 is then combined with the heavy condensate let down from Q-105 in the high pressure section and the mixture is then separated in the intermediate pressure hot condensing drum (Q-108). For reasons similar to those described in the high pressure section, a two-stage condensing system is used to first condense heavy hydrocarbons from the flashed vapor to a water washing step. Heavy condensate from Q-108 is let down to the low pressure hot condensing drum (Q-111). Vapor leaving Q-108 is water washed to dissolve sulfide and carbonate salts, partially condensed in M-107 and separated in Q-109 by a procedure similar to that described in the high pressure flash section. The washed hydrocarbons from Q-106 (light condensate from high pressure section) do not require reseparation of oil and water, hence, they are fed directly to the hydrocarbon section of the intermediate pressure cold condensing drum (Q-109). This has an advantage of reducing the drum volume required for phase separation. Sour water is pressured to the sour water accumulator (Q-114). Light condensate is let down in pressure to the low pressure cold condensing drum (Q-112) and non-condensable vapors are fed directly to the hydrogen recovery unit.

Low Pressure Level Flash Section

Reactor effluent slurry from the intermediate pressure level is flashed adiabatically to the low pressure slurry drum (Q-110). This low level flash section is used to remove light gases and to recover by adiabatic flashing, valuable hydrocarbon liquids from the coal ash slurry. Slurry from Q-110 is pumped to the recycle oil preparation section atmospheric flash drum (Q-701).

Flashed vapors are condensed and water-washed in a two step condensation procedure as in the high and intermediate pressure flash sections. The vapor stream from Q-110 is combined with the heavy condensate from Q-108, and the mixture is separated in Q-111. Heavy condensate from Q-111 is pumped to Section 600 for recovery and separation of hydrocarbon products. Vapor from Q-111 is water washed, cooled in air cooler M-108, and separated in the oil-water section of the low pressure cold condensing drum (Q-112). Light condensate from the intermediate pressure cold condensing drum (Q-109) is fed directly to the hydrocarbon section of Q-112. In drum Q-112, the vapor, hydrocarbon liquid, and water phases are separated.

Non-condensable vapors from Q-112, after removal of H_2S and CO_2 by amine absorption, are fed to the fuel gas header. Light hydrocarbon condensate from Q-112 is pumped to Section 600 for recovery and separation of hydrocarbon products while sour water is pressured to the sour water accumulator (Q-114).

Sour water from the three sections discussed above is let down to essentially atmospheric pressure and collected in the sour water accumulator (Q-114). Flash vapors from the sour water drum are vented to the fractionator overhead vent compressor where the vapors are mixed with off-gas from the fractionator, compressed and then sent to the gas treating plant for removal of sulfur compounds. Sour water is pumped to the sour water treating plant where ammonia and H_2S are stripped from the water.

SECTION 150, 250, 350, 450, & 550 - CATALYST HANDLING SYSTEM

One of the most significant advantages of the ebullated-bed reactor system is the ability to add and withdraw a portion of the catalyst without interrupting the operation of the plant. Catalyst addition and withdrawal are made batchwise on a once-per-day basis, thereby maintaining a constant level of catalyst performance.

SECTION 600 - DISTILLATE SEPARATION

INTRODUCTION

Section 600 is the distillate separation section of the CTSL Plant. It consists of a single train, designed for 8,400 Ton/Day dry coal. Feed comes to Section 600 from the five Slurry Preparation, Reaction, and Separation Trains, Sections 100 to 500, and also from Recycle Oil Preparation, Section 700. These streams are processed in the fractionator for product recovery. Overhead naphtha is stabilized before being transferred to storage. Distillates, as well as flush oil for the instrument bleeds and seal oil for the ebullating pumps are transferred to the product tank farms. The fractionator bottoms stream is recycled back to the reactor section where the material is used as a component of the slurry oil stream.

Drawing E-12-6-01 (Exhibit 2) is a process flow diagram for Section 600.

PROCESS FLOW

Light condensate products from the low pressure loop of Sections 100 to 500 are combined, preheated against hot heavy distillate in M-611A/B and phase separated in Desalter Preflash Drum (Q-601). The vapor stream is fed directly to the flash zone of the Fractionator. The liquid stream is pumped, under flow control, to the Desalter (S-601). Wash water to the Desalter consists of the condensate from the Scrubber Separator (Q-102 to 502) in Sections 100 to 500, and the sour water from Fractionator Overhead Receiver (Q-603). The wash stream is first preheated against the water effluent from the desalter in M-613, and is then contacted with the liquid hydrocarbon from Q-601 in a mixing valve upstream of the Desalter. In the Desalter, water is coalesced, decanted in the presence of an electric field, and drawn on interface level control. The desalter feed water flow rate is maintained by a split-range flow controller. Either the stripped sour water header provides make-up requirements, or excess water is sent to sour water treating. Salt-laden desalter effluent water is sent to waste water treatment (to prevent chloride buildup in the sour water recycle loop).

Treated hydrocarbons from the desalter are preheated by exchange with the light distillate pumparound and the heavy distillate pump around. The preheated stream is then mixed with heavy condensate from the low pressure loop of Sections 100 to 500, and the resulting two-phase mixture is separated in the Fractionator Preflash Drum (Q-602). Preflash drum vapors are fed directly to the fractionator flash zone. The liquid stream is pumped, under flow control, through the Fractionator Feed Heater (L-601). The heater outlet temperature is controlled by adjusting the rate of the fuel fired in the heater. Furnace efficiency is increased by superheating 50 psig and 150 psig steam in the convection section. By-passing saturated steam around the superheating coils is used to control the superheated steam outlet temperature.

The fractionator operating pressure is maintained by a pressure controller at the fractionator overhead receiver. The fractionator top temperature is controlled by adjusting the reflux flow rate to the top tray of the tower. In order to lift as much distillate-cut materials as possible, the Fractionator Feed Heater operates at a high heater outlet temperature with superheated stripping steam injected upstream of the heater. Steam on flow control is used to strip the fractionator bottoms. The fractionator bottoms stream is recycled back to the slurry mix tank to meet recycle slurry oil requirement.

Each side stream from the tower is a total liquid draw with the reflux pumped back to the tray below. The net heavy distillate product is stripped with superheated steam. The overhead stripped vapor is returned to the tower. The stripped heavy distillate product is pumped and cooled in the 200 psig and 75 psig steam generators, M-609 and M-610. The resulting stream from M-610 is further cooled against low pressure condensate in M-611A/B and then air-cooled. The cooled product is sent as needed to flush oil storage, with the balance of the heavy distillate product being combined with cooled light distillate product and pumped to distillate storage. The pumparound exchanger circuit is designed for maximum heat recovery from the heavy distillate pumparound. The heavy distillate pumparound is cooled by heat exchange against fractionator feed in M-603A/B, and further cooled in the 200 psig steam generator (M-604).

Light distillate is drawn from the tower. The light distillate product is then stripped with superheated steam. Cooled light distillate is combined with cooled heavy distillate product and pumped to distillate storage.

A neutralizing amine compound and a filming amine compound are injected through metering pumps into the fractionator overhead vapor line. The overhead vapor is condensed by air cooling in M-605 followed by a water-cooled trim condenser (M-606). Condenser effluent consisting of a liquid water phase, a liquid hydrocarbon phase, and a wet hydrocarbon vapor phase flow to the Fractionator Overhead Receiver (Q-603) where the three phases are separated. Sour water is drawn off the vessel boot and pumped by J-603 to the desalter. Condensed hydrocarbons are net naphtha product and tower reflux. The reflux to the tower is returned under flow control which is reset by top vapor temperature control. Naphtha product is pumped under flow control to the Naphtha Stabilizer (N-604).

Vapor from Q-603 flows to the vent gas compressor suction drum (Q-604) and is compressed by R-601 and sent to the naphtha stabilizer. Suction pressure is held constant in the overhead receiver by a compressor by-pass pressure controller. Flashed vapor from the sour water drums and wash water drums in Section 100 to 500 also flows to the suction of R-601 through Q-604.

The objective of the naphtha stabilizer is to remove C_4 and lighter materials from the naphtha product. Tower feed from Q-603 is preheated against the stabilizer bottoms. Overhead vapor is cooled and partially condensed in M-616. Vapor, water and hydrocarbon liquid are separated in the stabilizer overhead receiver (Q-605). Condensed liquid is used for tower reflux.

Vapor traffic in the stabilizer is maintained by sending the total liquid draw off the bottom tray through a once-through reboiler, (M-617). The reboiler uses steam as a heating medium. Tower bottoms are exchanged against incoming tower feed and then further cooled in the Stabilizer Trim Cooler (M-618) before being pressured on level control to stabilized naphtha storage.

SECTION 700 - RECYCLE OIL PREPARATION

PROCESS DESCRIPTION

The function of Section 700 in the CTSL Process is to provide a solids-free reactor recycle stream containing a sufficient quantity of residuum material such that the desired level of conversion is obtained in the reactor. The single-train recycle oil preparation section handles slurry from all five reactor trains, Sections 100 to 500.

Slurry received from Q-110 to Q-510 is mixed with superheated steam and flashed in the atmospheric flash drum (Q-701) to remove light ends from the slurry stream. The stripped light ends from Q-701 are returned as part of the vapor feed to N-601 fractionator. The slurry from Q-701 is then pumped to the Solid-Liquid Separation Unit (S-701). From the separation unit a solids-free recycle stream is pumped to the slurry mix tanks (Q-101 to Q-501). The net residuum and solids from the separation unit is fed to the partial oxidation gasifiers.

Drawing E12-7-01 (Exhibit 3) is a process flow diagram of Section 700.

DEASHING PROCESS PERFORMANCE

The CTSL Process design for this Illinois No. 6 coal study is based on providing a solids separation unit to prepare a solids-free recycle liquid. Operation with a solids-free recycle reduces the amount of recycle required to maintain a given oil-to-solids ratio in the coal feed slurry. It also allows higher residuum (975°F$^+$) concentration in the recycle liquid and as a result higher overall conversion of 975°F$^+$ liquid to distillable products.

The solids separation unit process performance for this design is based on the liquefaction process performance demonstrated with Wyodak coal in Run 227-27. The solid product rejected from the deashing unit (ash concentrate) contains the net yield of 975°F$^+$ liquid plus all of the unconverted coal and ash. Samples of topped separator bottoms from Run 227-27 were sent to Kerr-McGee and Sandwell Technologies for evaluation in their ROSE-SR(13) and CENTRAX(14) processes. A comparison of estimated performance, in terms of liquid recovery with these processes and that used in the design is as follows:

	Liquid Recovery, %
ROSE-SR	85
CENTRAX	100
Design	96.6

The ROSE-SR Process performance is estimated based on performance demonstrated with other feedstocks, primarily at Wilsonville. The CENTRAX Process performance is based on small-scale laboratory tests.

In actual practice, during bench scale operations, oils were extracted from filter cake using toluene solvent. With this procedure oil recovery from the cake was from 85 to 90% complete. CENTRAX projected 100% oil recovery based on laboratory scale tests and Kerr McGee estimated 85% recovery based on properties of the atmospheric bottoms well within the range of recovery used in this study with a net yield of 2.9% 975°F$^+$ residuum plus coal and ash for the heavy recycle case.

BLOCK FLOW DIAGRAM OF LIQUEFACTION

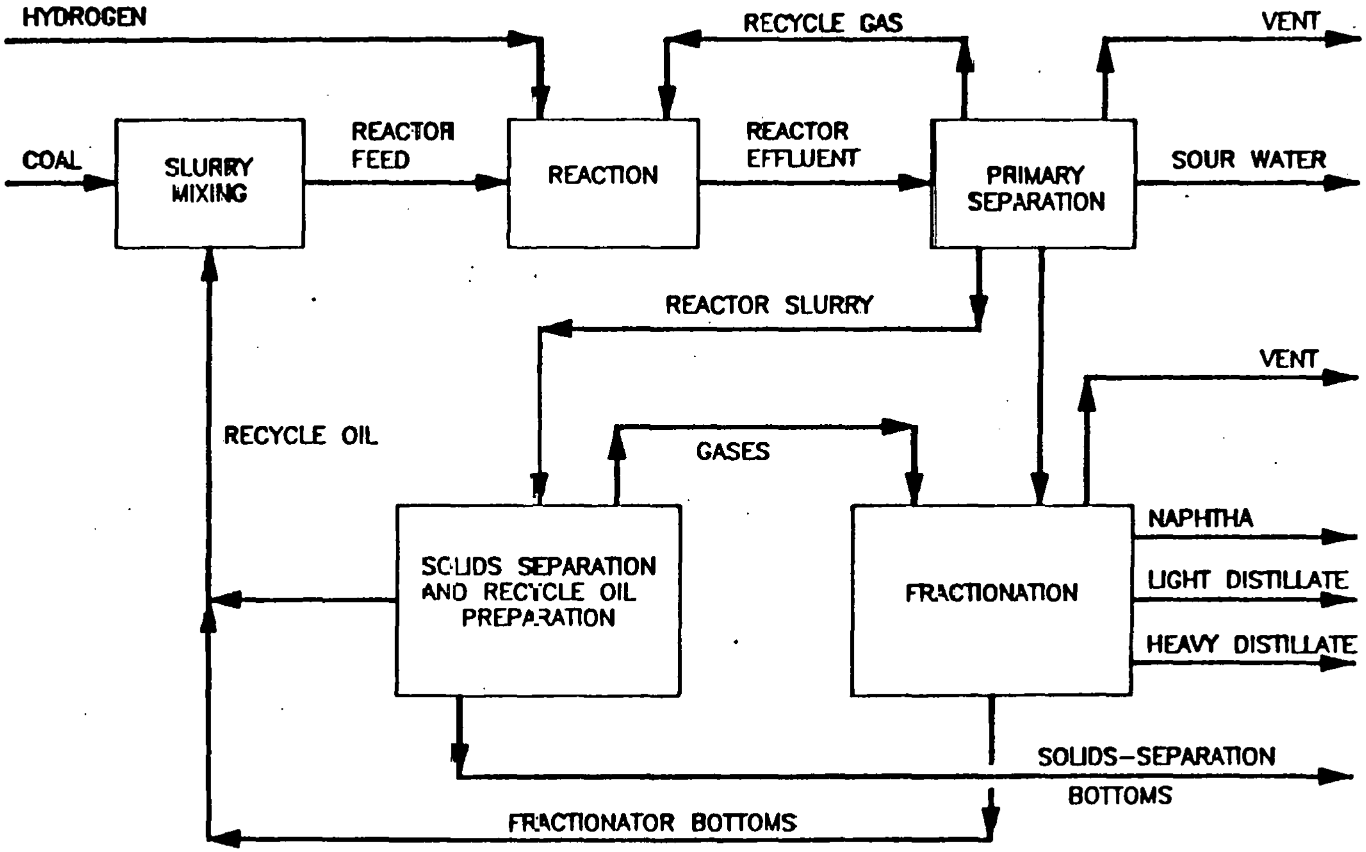


FIGURE 3

TABLE 13

AREA 200 - LIQUEFACTION

Material Balance Summary

DESCRIPTION	LBS/HR (DRY BASIS)
Coal Feed	700,000
Make-Up Hydrogen	52,408
Other Gases with Hydrogen	15,253
TOTAL	767,661
Stabilized Naphtha	136,326
Light Distillate	62,817
Heavy Distillate	286,377
Solids Separation Bottoms	81,000
DEA Absorber Bottoms (Net Pick-Up)	1,766
HPU Total Vent	72,385
Low Pressure Vent	15,326
Stabilizer Vent	12,261
Dissolved NH_3, H_2S, CO_2	25,413
Net Water Yield	73,990
TOTAL	767,661

TABLE 14

AREA 200 - LIQUEFACTION

Utilities Summary

Electric Power, Kw	63,811
Cooling Water, Gpm	4,520
Net Steam Produced	
600 Psig, Lb/Hr	60,767
200 Psig, Lb/Hr	10,095
Process Fuel, MMBtu/Hr	731

EQUIPMENT SPECIFICATIONS - PUMPS

ITEM No.	NUMBER REQUIRED OP/SPARE	SERVICE	DESIGN GPM	PSIG DISCHARGE/ SUCTION	TEMPERATURE °F	MATERIALS CASING	INTERNALS	BHP	REMARKS
J101-501A/B/C	10/5	Slurry Feed Booster	380	80/5	450	C.S. Fully Lined	25 Cr	25	
J102-502A/B	5/5	Eductor Liquid Circulating	1600	160/5	275	C.S. Fully Lined	25 Cr	594	
J103-503A/B	5/5	Scrubber Separator Steam Condensate	745	90/15	140	C.S.	C.S.	41	
J104-504	5	Scrubber Separator Skimmed Oil	50	75/0	140	C.S.	C.S.	7	
J105-505A/B/C	10/5	Reactor Feed	290	3275/50	450	Per Vendor Recommendation		653	Reciprocating
J106-506	5	Reactor Ebullating							
J107-507	5	Reactor Ebullating							
J108-508A/B	5/5	Low Pressure Flash Slurry Transfer	415	115/80	750	13 Cr	25 Cr Iron	12	
J109-509A/B	5/5	Heavy Distillate Feed	265	105/70	625	12 Cr	12 Cr	9	
J110-510A/B	5/5	Light Distillate Feed	195	95/60	120	C.S.	C.S.	7	
J111-511A/B	5/5	Process Water	130	3025/5	300	C.S.	C.S.	285	
J112-512A/B	5/5	Sour Water	185	75/0	115	12 Cr	12 Cr	15	
J113-513	5	Oily Slop	30	75/0	115	C.S.	C.S.	4	
J151-551A/B	5/5	Catalyst Transport Oil							
J152-552A/B	5/5	Circulating Oil							
J153-553	5/0	Catalyst Dump							
J154A/B	1/1	Circulating Oil							

TABLE 15

TABLE 15
(Continued)

EQUIPMENT SPECIFICATIONS - PUMPS

ITEM No.	NUMBER REQUIRED OP/SPARE	SERVICE	DESIGN GPM	PSIG DISCHARGE/ SUCTION	TEMPERATURE °F	MATERIALS CASING	MATERIALS INTERNALS	BHP	REMARKS
J601A/B	1/1	Fractionator Feed	1720	173/28	571	5% Cr	12 Cr	224	
J602A/B	1/1	Fractionator Overhead	910	126/11	100	C.S.	12 Cr	94	
J603A/B	1/1	Sour Water	70	131/11	100	316 SS	316 SS	16	
J604A/B	1/1	Light Distillate P/A	2400	123/48	456	C.S.	12 Cr	161	
J605A/B	1/1	Heavy Distillate P/A	2850	116/41	585	C.S.	12 Cr	192	
J606A/B	1/1	Fractionator Bottoms	570	76/26	695	12 Cr	12 Cr	26	
J607A/B	1/1	Light Distillate	210	111/31	429	C.S.	12 Cr	19	
J608A/B	1/1	Heavy Distillate	1250	130/25	556	C.S.	12 Cr	118	
J609A/B	1/1	Desalter Feed	920	97/27	250	C.S.	12 Cr	59	
J610A/B	1/1	Stabilizer Reflux	67	104/68	100	C.S.	12 Cr	5	
J-701A/B	1/1	Atmospheric Flash Slurry Transfer	1915	67/12	736	13 Cr	25 Cr		

TABLE 16

EQUIPMENT SPECIFICATIONS - REACTORS

ITEM No.	SERVICE	SIZE I.D.	SIZE T.T.	DESIGN TEMP., °F	DESIGN PRESSURE, PSIG	MATERIALS OF CONSTRUCTION	REMARKS
K101-501 and K-102-502	H-Coal® Reactor	14'0"	53'0"				

NOTES:

1. Ten (10) identical reactors required.

TABLE 17

EQUIPMENT SPECIFICATIONS - FIRED HEATERS

ITEM No.	NUMBER REQUIRED	SERVICE	DESIGN DUTY, MMBtu/Hr	OUTLET TEMP., °F	OUTLET PRESSURE, PSIG	MATERIALS OF CONSTRUCTION	REMARKS
L101-501	5	Slurry Feed	20.5	584	3065	347 S.S. or Incoloy 801	
L102-502	5	Hydrogen	24.0	950	3065	347 S.S.	
L-601	1	Fractionator Feed Heater	120.0	750	22	9 Cr-1 Mo Tube Radiant; C.S. Tube Convection	Convection coil to superheat 50 psig and 150 psig steam. Design Duty 6.1 MMBtu/Hr.

TABLE 18

EQUIPMENT SPECIFICATIONS - HEAT EXCHANGERS

ITEM No.	NUMBER REQUIRED OP/SPARE	SERVICE	TYPE	DUTY, MMBtu/Hr	AREA Ft^2	DESIGN - SHELL/TUBE PRESSURE, PSIG	DESIGN - SHELL/TUBE MATERIALS OF CONSTRUCTION	REMARKS
M101-501	5/0	Eductor Oil Cooler	Air Cooler	2.2	2250*	100	C.S.	
M102-502	5/0	Scrubber Separator P/A Exchanger	S & T	5.0	950	100/100	C.S./C.S.	
M103	1/0	Make-up Hydrogen Bypass Cooler	Air Cooler	17.3	3250*	600	C.S.	
M104-1 to M104-6	4/2	Make-up Hydrogen Interstage Cooler	Air Cooler	6.6	1500*	1800	C.S.	
M105-505A/B	5/0	Reactor Effluent Vapor/Hydrogen Feed						
M106-506	5/0	Reactor Effluent Cooler						
M107-507	5/0	Intermediate Pressure Vapor	Air Cooler	3.6	700*	875	304S.S.	
M108-508	5/0	Low Pressure Vapor	Air Cooler	3.1	700*	100& Full Vac.	304S.S.	
M109-509	5/0	Eductor Oil Steam Generator	Kettle	2.3	130	300/100	C.S./11-13Cr	
M110-510	5/0	Eductor Oil Scrubber Bottom Purge	S & T	1.6	230	50& Full Vac./100	C.S./5Cr-1/2Mo	
M151-551	5/0	Circulating Oil	Air Cooler	1.2	1500*	150	C.S.	
M601A/B	2/0	Light Distillate/ Fractionator Feed	S & T	23.9	7100	150/150	C.S./ 11-13Cr	
M602A/B	1/0	Light Distillate P/A Trim	Air Cooler	6.4	1050*	150	11-13Cr	Fan BHP 15
M603A/B	2/0	Heavy Distillate P/A Fractionator Feed	S & T	37.0	8600	150/150	1 1/4Cr-1/2Mo/ 5Cr-1/2Mo	
M604	1/0	Heavy Distillate P/A 200 Psig Steam	Kettle	35.7	4550	300/150	C.S./5Cr-1/2Mo	
M605	1/0	Fractionator Overhead Condenser	Air Cooler	95.3	18420*	150& Full Vac.	C.S.	Fan BHP 313
M605	1/0	Fractionator Overhead Trim Cooler	S & T	7.3	4200	150& Full Vac./100	C.S./C.S.	
M607	1/0	Light Distillate/ 75 Psig Steam	Kettle	3.2	700	150/150	C.S./11-13Cr	

*Bare Tube Surface Area

TABLE 10 (Continued)

EQUIPMENT SPECIFICATIONS - HEAT EXCHANGERS

ITEM No.	NUMBER REQUIRED OP/SPARE	SERVICE	TYPE	DUTY, MMBtu/Hr	AREA Ft^2	DESIGN - SHELL/TUBE PRESSURE, PSIG	DESIGN - SHELL/TUBE MATERIALS OF CONSTRUCTION	REMARKS
M608	1/0	Light Distillate Trim Cooler	Air Cooler	6.8	4800*	150	11-13Cr	Fan BHP 71
M609	1/0	Heavy Distillate/ 200 Psig Steam	Kettle	24.0	2750	300/150	C.S./5Cr-1/2Mo	
M610	1/0	Heavy Distillate/ 75 Psig Steam	Kettle	19.4	3150	150/150	C.S./C.S.	
M611	1/0	Heavy Distillate/ LP Condensate	S & T	23.4	6110	150/150	C.S./C.S.	
M612	1/0	Heavy Distillate/Trim	Air Cooler	19.7	23500*	150	C.S.	Fan BHP 352
M613	1/0	Desalter Wash/Effluent	S & T	1.7	260	150/150	C.S./C.S.	
M615	1/0	Stabilizer Feed Bottoms	S & T	10.0	1800	150/100	C.S./C.S.	
M616	1/0	Stabilizer Overhead Condenser	S & T	3.0	1850	100/75	C.S./C.S.	
M617	1/0	Stabilizer Reboiler	S & T	11.6	4500	100/225	C.S./C.S.	
M618	1/0	Stabilizer Naphtha Trim Cooler	S & T	9.6	2700	100/75	C.S./C.S.	

*Bare Tube Surface Area

TABLE 19

EQUIPMENT SPECIFICATIONS - TOWERS

ITEM No.	NUMBER REQUIRED OP/SPARE	SERVICE	SIZE I.D.	SIZE T.T.	DESIGN TEMP., °F	DESIGN PRESSURE, PSIG	NUMBER OF TRAYS	MATERIALS OF CONSTRUCTION	REMARKS
N601	1/0	Fractionator	Top 12'0"	17'6"	650	50 & Full Vac.	5	C.S. 317SS Clad	
			Transition	2'6"				C.S. 317SS Clad	
			Middle 15'	58'6"			15	C.S. 317SS Clad (23'6") C.S. 410SS Clad (35')	
			Transition	8'0"					
			Bottom 5'	24'	775		4	C.S. 410SS Clad	
N602	1/0	Light Distillate Stripper	3"0"	30'	650	50 & Full Vac.	4	C.S. 317SS Clad	
N603	1/0	Heavy Distillate Stripper	6'6"	34'	650	50 & Full Vac.	4	C.S. 410SS Clad	
N604	1/0	Naphtha Stabilizer	Top 3'	15'6"	300	100	7	C.S.	
			Transition 3'6"					C.S.	
			Bottom 6'0"	55'	450	100	17	C.S.	

TABLE 20

EQUIPMENT SPECIFICATIONS - VESSELS

ITEM No.	NUMBER REQUIRED OP/SPARE	SERVICE	HORIZ. OR VERT.	SIZE I.D.	SIZE T.T.	DESIGN TEMP., °F	DESIGN PRESSURE, PSIG	MATERIALS OF CONSTRUCTION	REMARKS
Q101-501	5/0	Slurry Mix Tank	V	11'0	29'6"	650	50 & Full Vac.	C.S.	Hot Insulation Required
Q102-502	5/0	Scrubber Separator	V	Upper 3'6" Lower 6'0" Transition 2'6"	21'6" 10'0"	650	50 & Full Vac.	C.S.	Upper sections to include 10'0" Packed Bed Section of 2" metal Pall Rings
Q103-503	5/0	Skimmed Oil Accumulator	V	1'6"	4'0"	250	50 & Full Vac.	C.S.	
Q104-504	5/0	Reactor Effluent Separator							
Q105-505	5/0	Reactor Effluent Hot Condensing Drum							
Q106-506	5/0	Reactor Effluent Cold Condensing Drum							
Q107-507	5/0	Intermediate Pressure Slurry Drum	V	4'0"	14'0"	850	875	1¼Cr-½Mo 304S.S. Internals	Hot insulation required.
Q108-508	5/0	Intermediate Pressure Hot Condensing Drum	V	5'0"	14'0"	750	875	1¼Cr-½Mo 304S.S. Internals	Hot insulation required.
Q109-509	5/0	Intermediate Pressure Cold Condensing Drum	H	6'0"	19'0"	450	875	C.S. 304S.S. Internals	Gunnite lined. Nozzles monel clad
Q110-510	5/0	Low Pressure Slurry Drum	V	6'6"	20'6"	850	100 & Full Vac.	1¼Cr-½Mo 304S.S. Internals	Hot insulation required.
Q111-511	5/0	Low Pressure Hot Condensing Drum	V	6'0"	16'6"	750	100 & Full Vac.	1¼Cr-½Mo 304S.S. Internals	Hot insulation required.
Q112-512	5/0	Low Pressure Cold Condensing Drum	H	6'6"	21'6"	450	100 & Full Vac.	C.S. 304S.S. Internals	Gunnite lined.
Q113-513	5/0	Recycle Compressor KO Drum	V	5'0"	10'0"	450	3300	C.S.	
Q114-514	5/0	Sour Water Accumulator	H	7'0"	21'0"	300	50 & Full Vac.	C.S.	Gunnite lined.
Q115-515	5/0	Wash Water Receiver	V	6'0"	15'0"	400	50 & Full Vac.	C.S.	
Q116-1-3	3/0	Make-up Hydrogen Suction KO Drum	V	10'6"	9'6"	450	600	C.S.	
Q117-1-6	6/0	Make-up Hydrogen Interstage KO Drum	V	4'0"	9'6"	450	1800	C.S.	

TABLE 20 (Continued)

EQUIPMENT SPECIFICATIONS - VESSELS

ITEM No.	NUMBER REQUIRED OP/SPARE	SERVICE	HORIZ. OR VERT.	SIZE I.D.	SIZE T.T.	DESIGN TEMP., °F	DESIGN PRESSURE, PSIG	MATERIALS OF CONSTRUCTION	REMARKS
Q151	1/0	Fresh Catalyst Holding Tank							
Q152-552	5/0	Catalyst Hopper							
Q153-553	5/0	Catalyst Addition Drum							
Q154-554	5/0	Catalyst Transport Oil Drum							
Q155-555	5/0	Catalyst Withdrawal Drum							
Q156-556	5/0	Catalyst Withdrawal Vapor Separator							
Q157-557	5/0	Catalyst Cooling Drum							
Q158-558	5/0	Circulating Oil Drum							
Q159	1/0	Spent Catalyst Holding Tank							
Q160-560	5/0	Catalyst Dump Tank							
Q161-561	5/0	Catalyst Dump Tank							
Q601	1/0	Desalter Preflash Drum	H	8'6"	23'	650	50 & Full Vac.	C.S.	Hot insulation required
Q602	1/0	Fractionator Preflash Drum	H	10'6"	29'6"	650	50 & Full Vac.	C.S. 410 SS Clad	Hot insulation required
Q603	1/0	Fractionator Overhead Receiver	H	12'	31'6"	400	50 & Full Vac.	C.S.	
Q604	1/0	Compressor Suction Drum	V	3'6"	8'6"	350	50	C.S.	
Q605	1/0	Stabilizer Overhead Receiver	H	5'6"	14'6"	300	100	C.S.	
Q701	1/0	Atmospheric Flash Drum	V	10'6"	33'	850	50 & Full Vac.	C.S. 410 SS Internals and Cladding	Hot Insulation Required

EQUIPMENT SPECIFICATIONS - COMPRESSORS

ITEM No.	R101-1-6A/B	R102-502	R601A/B
NUMBER REQUIRED OPERATING/SPARE	4/2	5/0	1/1
SERVICE	Make-up Hydrogen	Recycle Hydrogen	Vent Gas
TYPE	Reciprocating (2-Stage)	Centrifugal	Reciprocating
DESIGN, MMSCFD	105	131	1.7
TEMPERATURE, °F (Suction)	30	130	100
PRESSURE, PSIG			
Suction	535	2950	3
Discharge	3275	3225	85
MW of GAS	3.30	8.87	43.2
DRIVE	Electric	Electric	Electric
BHP/MACHINE	12,050	970	162

ITEMS S-101 to S-501

SLURRY MIX TANK EDUCTOR

(Five Items Required)

DESCRIPTION

Liquid eductor using hydrocarbon motive liquid to remove vapors and entrained coal dust particles escaping from the slurry mix tank. Vendor shall design provisions for a continuous wash at suction area to prevent bridging and plugging due to presence of coal dust.

PROCESS DUTY

Suction

Pressure	0.5 psig (maximum)
Temperature	450°F

Eductor Discharge Pressure 1.0 psig (minimum)
(Vendor to provide his maximum design discharge pressure for the given suction load and motive liquid.)

INLET STREAM TO EDUCTOR

	M.W.	Lbs/Hr
Water Vapor	18	4,622
Hydrocarbons	279	1,229
Total		5,851

Inlet stream will also contain coal dust (estimated 200 lbs/hr.).

MOTIVE LIQUID

4.3°API coal-derived hydrocarbon
4,100 GPM at 100 psig and 275°F at eductor nozzle
Motive liquid may contain up to 1.0 W % coal dust particles
(About 100 micron size)

MATERIAL OF CONSTRUCTION

Eductor Nozzle - Stainless Steel
Suction Chamber/Diffuser - Cast iron or fabricated steel.

S-102
HYDROGEN PURIFICATION UNIT

DESCRIPTION Provide a hydrogen purification unit for the recovery of a high purity hydrogen stream. Vendor to include an amine absorption unit for the removal of H_2S and CO_2 compounds present in the feed stream.

FEED STREAM		
	Temperature	120°F
	Pressure	700 psig

COMPONENT	MOLS/HR	LBS/HR	MOL. FRACTION
H_2	10351.2	20868	0.7135
N_2	802.8	22477	0.0553
CO	885.0	24781	0.0610
H_2S	50.1	1708	0.0035
NH_3	2.5	42	0.0002
H_2O	26.6	478	0.0018
CO_2	1.4	59	0.0001
CH_4	1397.3	22356	0.0963
C_2H_6	501.7	15102	0.0346
C_3H_8	308.9	13622	0.0213
NC_4	106.9	6212	0.0074
$C5^+$	70.0	5866	0.0050
TOTAL	14504.4	133571	1.0000

PRODUCT SPECIFICATION

Hydrogen Concentration, Min.	88 V % Min.
Temperature	120°F Max.
Pressure	580 psig min.
Hydrogen Recovery	99.3 % min.

MECHANICAL SPECIFICATION

A. Vessels to be designed and stamped in accordance with ASME code.
Electrical installation for Class I. Div. I area.
Exchangers to meet TEMA R Specifications
Vessels and exchangers to be specified with 1/8" corrosion allowance.

B. Fuel gas header pressure = 50 psig

S-151 TO S-551
CATALYST HANDLING EJECTOR

(Five Required)

TYPE:	Single stage vacuum jet complete with surface type aftercondensor.		
DUTY:	Ejector system to be capable of evacuating a system volume of 375 ft^3, initially at atmospheric pressure and ambient tempemperature to a final pressure of 100mm Hg. abs. within 15 minutes. System is initially filled with air. Discharge pressure of jet is 1 psig.		
UTILITIES:	Cooling water	-	85°F, 30°F maximum rise
	Steam	-	150 psig saturated, may contain up to 1.0 W % moisture.
MATERIAL OF CONSTRUCTION:	Manufacturer's standard.		

S-152 TO S-552

CIRCULATING OIL STRAINER

(Five Required)

TYPE:	In-line basket type mesh screen strainer.
FLUID STRAINED:	4.3°API Vacuum Gas Oil 24,500 Lbs/Hr
OPERATING TEMP:	250°F
SPECIFIC GRAVITY AT 250°F:	0.98
PARTICLES REMOVED:	Catalyst fines, coal fines, coke.
PARTICLES SIZE TO BE REMOVED:	0.032" and larger

TABLE 26

S-153

MAKE-UP OIL STRAINER

(One Unit Required)

TYPE:	In-line basket type mesh screen
FLUID STRAINED:	4.3°API Oil 10,600 Lb/Hr
OPERATING TEMPERATURE:	695°F Maximum
SPECIFIC GRAVITY: AT 695°F	0.84
PARTICLES REMOVED:	Coal and ash fines
PARTICLES SIZE REMOVED:	<5 Micron

S-601 - DESALTER

ELECTRICAL DESALTER DUTY SPECIFICATION

HYDROCARBON FEED	
Boiling Range	C_4 to 950°F
UOP "K"	11.1
Gravity	32.2 °API (0.78 s.g. @ 245°F)
Viscosity	0.65 cp @ 245°F
Vapor Pressure	37 PSIA @ 245°F
Moisture	Negligible
Solids	0.1 W % (Max.) Coal Ash (10 to 100 Micron Particles)
Salt	Ammonium Chloride (NH_4Cl) 5 to 10 lb per 1000 Bbl
PERFORMANCE	
Design Throughput	28,600 Bbl/Day
Operating Conditions	65 Psig (Outlet); 245°F
Design Conditions	150 Psig; 650°F
Design Duty	Reduce the hydrocarbon stream salt concentration to 1 Lb/1000 Bbl.
WASH WATER	
Available Rate	1950 Bbl/Day
Source	Sour water from distillate fractionator.
ALLOWABLE HYDROCARBON ΔP:	10 Psi (Inlet to outlet)

EMULSION REMOVAL

Provide three (3) perforated collection headers for manual emulsion drawoff. Locate these headers horizontally at 9", 18" and 27" above the desalter inlet feed distributor. Each collection header is to have a separate draw-off nozzle.

S-602

NEUTRALIZING AMINE INJECTION UNIT

SERVICE

To provide a preassembled, skid mounted equipment package for the purpose of storing and metering a controlled dosage of a liquid corrosion inhibitor to the overhead vapor line of the atmospheric fractionator, N-601.

COMPONENTS

Two (2) metering pumps capable of delivering 2.0 gal/hr liquid corrosion inhibitor at 100 psig (atm. suction). Standard screw adjustment required from zero to full stroke.

Materials: Pump Body = C.S.
Plungers or Diaphrams - S.S.

One (1) atmospheric storage tank adequate for seven (7) days' inventory of liquid corrosion inhibitor (250 gal. min.)

Materials: C.S. or FRP

OTHER EQUIPMENT

Vendor to advise any requirements for dilution, dispersion, or agitation. Vendor to provide all interconnecting piping and instrumentation (including alarm connections to purchaser's system and calibrate gauge glass).

CORROSION INHIBITOR

Vendor to recommend a specific neutralizing amine (morpholine, cyclohexylamine, etc.) for a fractionator overhead dew-point vapor which is cooled from 325 to 100°F.

	VAPOR AT		CONDENSED (100°F)	
POUNDS/HOUR	325°F	100°F	HYDROCARBONS	WATER
Total	315,410	7,250	279,640	28,520
H_2S	563	306		
CO_2	5	4		
NH_3	42	17		
Chlorides	Trace			

S-603

FILMING AMINE INJECTION

SERVICE

To provide a preassembled, skid mounted equipment package for the purpose of storing and metering a controlled dosage of a liquid filming-type amine to the overhead vapor line of the atmospheric fractionator, N-601.

COMPONENTS

Two (2) metering pumps capable of delivering 0.5 gal/hr liquid filming amine at 100 psig (atm. suction). Standard screw adjustment from zero to full stroke.

Materials: Pump Body = Carbon Steel
Plungers or Diaphrams - Stainless Steel

One (1) atmospheric storage tank adequate for seven (7) days' inventory of liquid filming-type amine (250 gal. min.)

Materials: Carbon steel or FRP

INSTRUMENTATION

Vendor to provide alarm connections to purchaser's system. The storage tank shall be equipped with a calibrated gauge glass. Piping shall be arranged such that liquid can be pumped either from the tank or directly from the calibrated gauge glass.

FILMING AMINE

Vendor to recommend a specific filming-type amine (octadecylamine, etc.) and injection rate for a fractionator overhead dew-point vapor which is cooled and partially condensed from 325 to 100°F. (A dosage rate of 5 ppm (weight) corresponds to 0.22 gal/hr.)

ITEM S-701

SOLID-LIQUID SEPARATOR (DEASHING) UNIT

A solid-liquid separation (deashing) unit is provided to recover a solids-free, residuum containing (975°F$^+$) liquid stream for liquefaction. In the bench-scale Demonstration Run 227-47(I-27) topped separator bottoms were filtered and the filtered liquid was used as recycle. At commercial catalyst ages there is a small net yield of residual (975°F$^+$) oil. Deashing unit performance was set such that the ash concentrate product contained this net yield of residual oil.

The overall material balance for the solids separation unit is as follows:

FEED, LB/HR		
Liquid	635,435	(-7.8 °API Coal Derived Oil)
Solids	59,780	(32.4 W % Unconverted Coal, 67.6 W % Ash)
	695,215	

PRODUCT, Lb/Hr	
Recycle Liquid	614,215
Ash Concentrate	81,000
	695,215

EXHIBIT 1

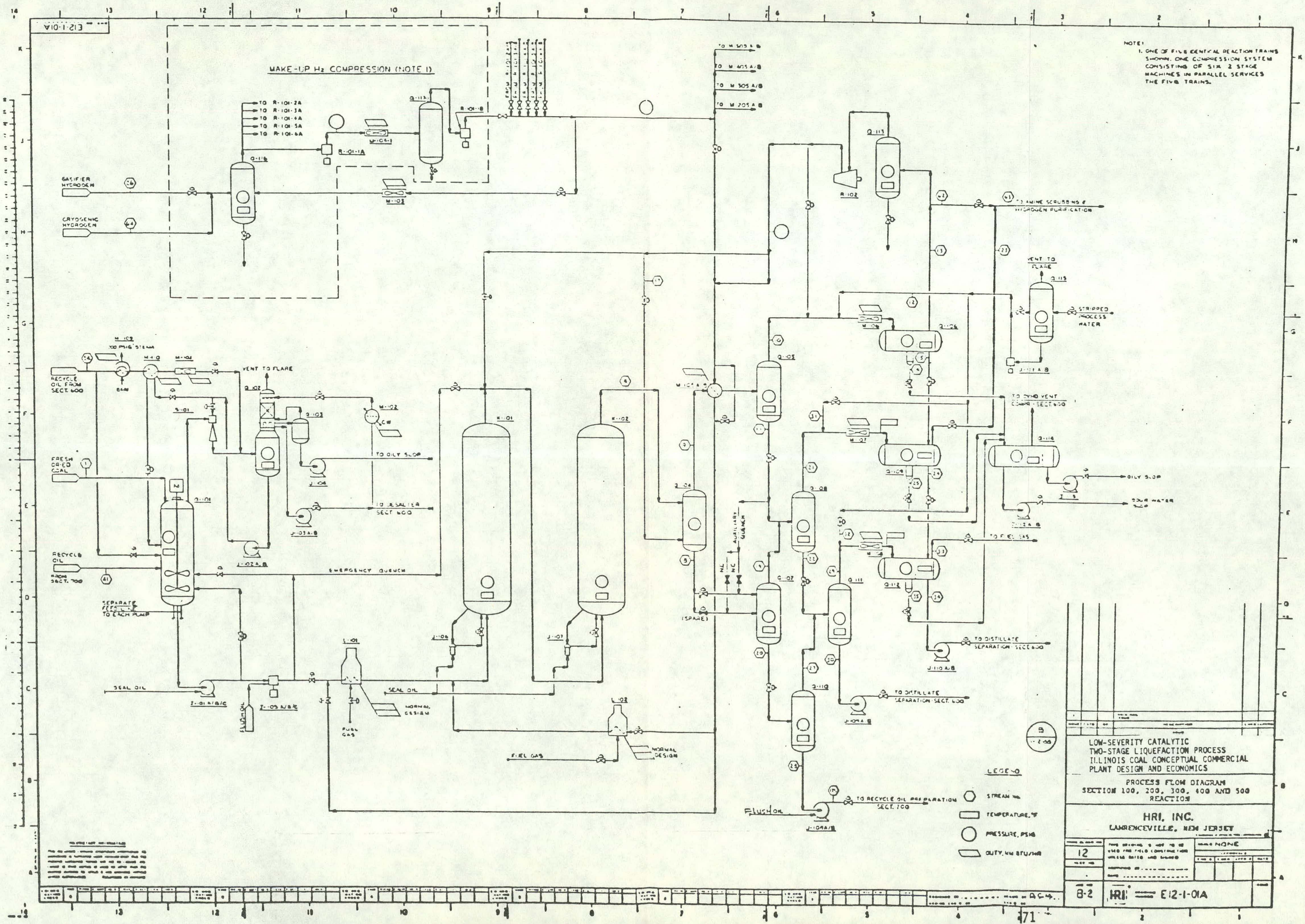
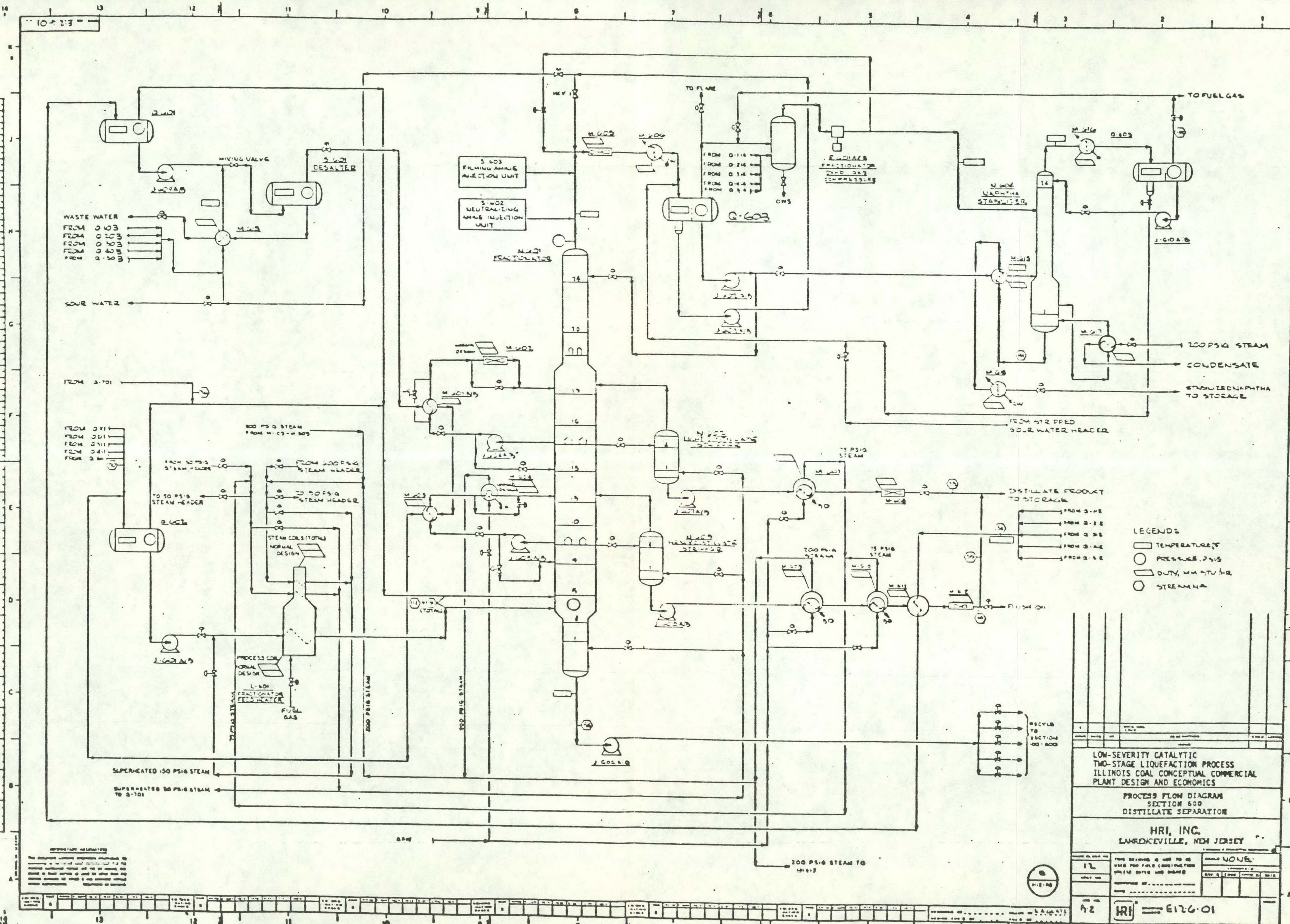

NOTE:
1. ONE OF FIVE IDENTICAL REACTION TRAINS SHOWN. ONE COMPRESSION SYSTEM CONSISTING OF SIX 2 STAGE MACHINES IN PARALLEL SERVICES THE FIVE TRAINS.
MAKE-UP H2 COMPRESSION (NOTE 1)
GASIFIER HYDROGEN
CRYOGENIC HYDROGEN
RECYCLE OIL FROM SECT. 600
FRESH DRIED COAL
RECYCLE OIL FROM SECT. 700
VENT TO FLARE
TO OILY SLOP
TO DESALTER SECT. 600
EMERGENCY QUENCH
SEAL OIL
FLUSH OIL
FUEL GAS
NORMAL DESIGN
TO AMINE SCRUBBING & HYDROGEN PURIFICATION
STRIPPED PROCESS WATER
TO DHO VENT COMPR - SECT. 600
OILY SLOP
SOUR WATER
TO FUEL GAS
TO DISTILLATE SEPARATION SECT. 600
TO RECYCLE OIL PREPARATION SECT. 700
LEGEND
STREAM No.
TEMPERATURE, °F
PRESSURE, PSIG
DUTY, MM BTU/HR
LOW-SEVERITY CATALYTIC
TWO-STAGE LIQUEFACTION PROCESS
ILLINOIS COAL CONCEPTUAL COMMERCIAL
PLANT DESIGN AND ECONOMICS
PROCESS FLOW DIAGRAM
SECTION 100, 200, 300, 400 AND 500
REACTION
HRI, INC.
LAWRENCEVILLE, NEW JERSEY
NONE
12
B-2
HRI E12-1-01A

EXHIBIT 2


LOW-SEVERITY CATALYTIC
TWO-STAGE LIQUEFACTION PROCESS
ILLINOIS COAL CONCEPTUAL COMMERCIAL
PLANT DESIGN AND ECONOMICS
PROCESS FLOW DIAGRAM
SECTION 600
DISTILLATE SEPARATION
HRI, INC.
LAWRENCEVILLE, NEW JERSEY
NONE
E126-01
TO FUEL GAS
TO FLARE
DESALTER
MIXING VALVE
WASTE WATER
SOUR WATER
Q-603
200 PSIG STEAM
CONDENSATE
DISTILLATE PRODUCT TO STORAGE
FLUSH OIL
LEGEND:
TEMPERATURE, °F
PRESSURE, PSIG
DUTY, MM BTU/HR
STREAM No.
SUPERHEATED 150 PSIG STEAM
FUEL GAS

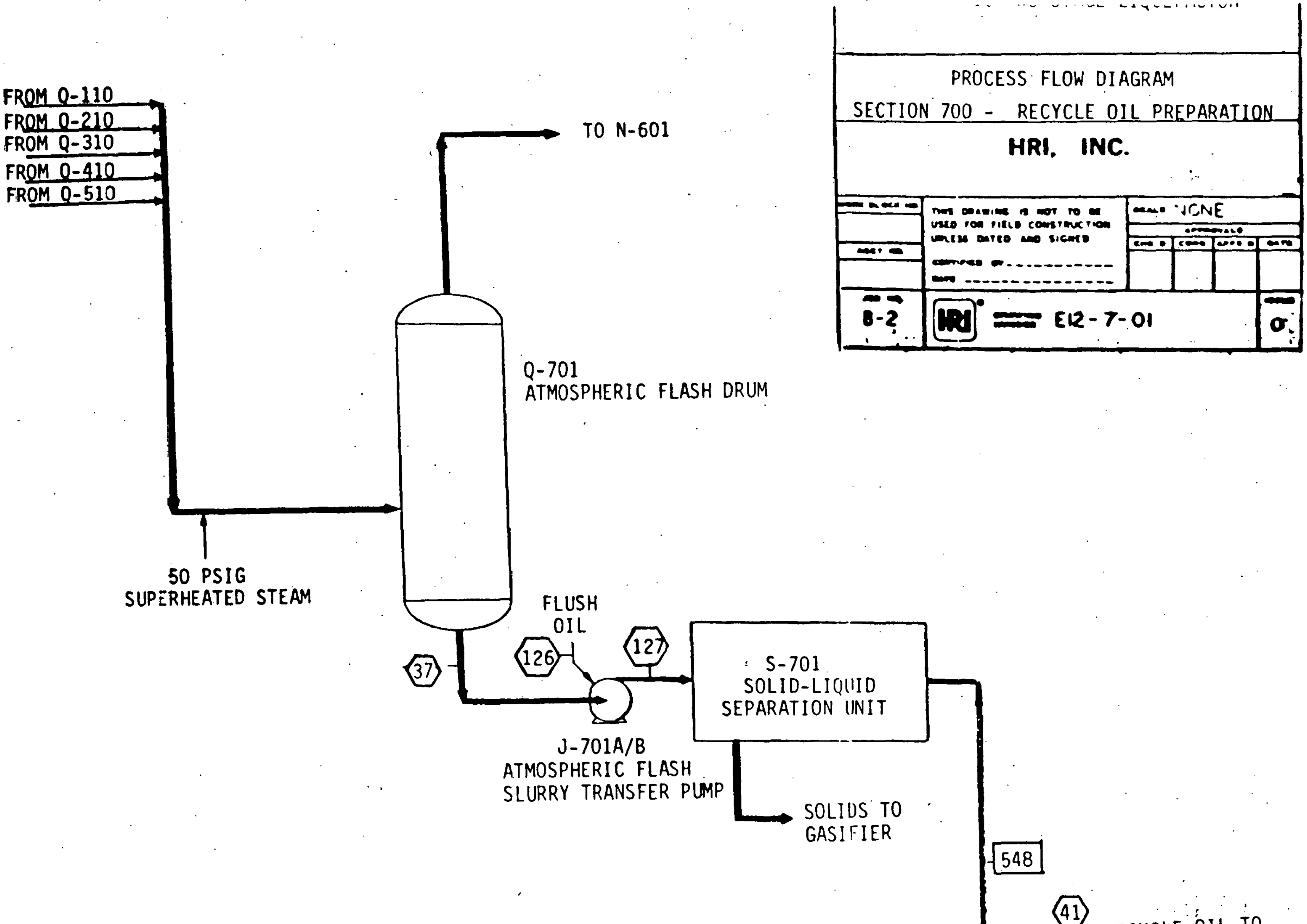

PROCESS FLOW DIAGRAM
SECTION 700 - RECYCLE OIL PREPARATION
HRI, INC.
THIS DRAWING IS NOT TO BE USED FOR FIELD CONSTRUCTION UNLESS DATED AND SIGNED
NONE
8-2
E12-7-01
FROM Q-110
FROM Q-210
FROM Q-310
FROM Q-410
FROM Q-510
TO N-601
Q-701
ATMOSPHERIC FLASH DRUM
50 PSIG
SUPERHEATED STEAM
FLUSH
OIL
126
127
37
S-701
SOLID-LIQUID
SEPARATION UNIT
J-701A/B
ATMOSPHERIC FLASH
SLURRY TRANSFER PUMP
SOLIDS TO
GASIFIER
548
41
RECYCLE OIL TO
Q-101-501

EXHIBIT 3

AREA 300 - PARTIAL OXIDATION

Partial oxidation is used to gasify the deashing process ash concentrate as well as the high-ash concentrate from deep-cleaning in Area 100. Gasification takes place at elevated temperatures and pressures in the presence of steam and oxygen. The ash concentrate is fed to the gasifier slurried with water. Ash is withdrawn from the gasification vessel. The primary products of gasification are H_2, CO, CO_2, and H_2O. Carbon monoxide in the product gas is shifted to hydrogen via the water gas reaction. Acid gases are removed using conventional technology.

Steam and oxygen requirements, as well as hydrogen yields and selectivities, are based on the extensive partial oxidation testing of liquefaction residues done by Texaco under contract to DOE.[6] HRI has analyzed this data and developed performance correlations versus feedstock qualities. The partial oxidation feedstock qualities for this conceptual design are as follows:

Feedstock Composition, W % on Dry Coal	Vacuum Bottoms	Coal
975°F$^-$	0.14	
975°F$^+$	2.89	
Unconverted Coal	2.77	77.00
Ash	5.77	23.00
Total	11.57	100.00
Percent Solids	73.8	100.00
HHV, Btu/Lb (M.A.F. Basis)	9,944	13,988

Investment and utility estimates are based on information previously provided to HRI by Texaco as well as other published information.

Material balance and utility summaries are provided in Tables 31 and 32.

TABLE 31

AREA 300 - PARTIAL OXIDATION

MATERIAL BALANCE SUMMARY

FEEDS, LB/HR	
Solids Separation Bottoms	81,000
Coal (1)	367,110
Light Gases	2,183
Oxygen	313,135
Steam (2)	410,940
TOTAL	1,174,368
PRODUCTS, LB/HR	
Make-up Hydrogen (3)	62,745
Ash	124,810
SO_2	21,900
Other Gases	964,913
TOTAL	1,174,368

(1) Dry Coal Basis. Fed to gasifier as 30% water slurry.
(2) Includes water to slurry coal feed.
(3) 98 V % hydrogen.

TABLE 32

AREA 300 - PARTIAL OXIDATION

UTILITIES SUMMARY

Electric Power, Kw	12,852
Cooling Water, Gpm	29,847
Net Steam Produced	
75 Psig, Lb/Hr	600,505
Raw Water, Gpm	337
Process Fuel, MMBtu/Hr	28

AREA 400 - OXYGEN PLANT

The oxygen plant produces high purity (99.5%) oxygen for use in gasification of deashing ash concentrate. Oxygen compression facilities are included. Excess nitrogen is available for export. No credit for nitrogen sales is taken in this evaluation. The technology used is conventional.

Material balance and utility summaries are provided in Tables 33 and 34.

TABLE 33

AREA 400 - OXYGEN

MATERIAL BALANCE SUMMARY

Products, Lb/Hr (99.5 V % Oxygen)	313,135

TABLE 34

AREA 400 - OXYGEN

UTILITIES SUMMARY

Turbine Power, Kw	65,937
Cooling Water, Gpm	29,299

AREA 500 - PRODUCT TREATING

Product treating contains four major areas.

1) Light ends recovery using conventional refining technology. C_4s are separated for blending to gasoline and for sale as LPG The lighter gases are used as plant fuel.

2) Acid gases (CO_2 and H_2S) are removed from product gases using conventional technology.

3) Collected sour water from the plant is stripped with steam. H_2S and NH_3 are recovered. NH_3 is recovered for by-product sales. The waste water is further treated in a phenolsolvan unit to remove dissolved phenols. No credit is taken for the phenols recovered.

4) Elemental sulfur is recovered from H_2S in a Claus-type sulfur plant.

Material balance and utility summaries are provided in Tables 35 and 36.

TABLE 35

AREA 500 - PRODUCT TREATING

MATERIAL BALANCE SUMMARY

Acid Gas Removal, Lb/Hr	
H_2S	2,784
CO_2	82
Sour Water Stripping, Lb/Hr	
Sour Water	428,286
CO_2 Removed	1,087
H_2S Removed	14,824
NH_3 Recovered	11,946
Stripped Sour Water	400,429
TOTAL	428,286
Sulfur Plant, Lb/Hr	
Sulfur Recovered	27,390
Light Ends Plant, Lb/Hr	
Process Fuel	80,098
C_4 to Blending	11,186
C_5^+ to Upgrading	9,195
LPG	2,199
Partial Oxidation Feed	2,183

TABLE 36

AREA 500 - PRODUCT TREATING

UTILIITIES SUMMARY

Electric Power, Kw	2,398
Cooling Water, Gpm	4,830
Net Steam Consumed	
75 Psig, Lb/Hr	110,663
600 Psig, Lb/Hr	68,511
Process Fuel, MMBtu/Hr	4

AREA 600 - PRODUCT UPGRADING

The net liquid product is hydrotreated at high severity over a fixed bed of catalyst. The hydrotreated product is fractionated to recover light naphtha (C_5-180°F), heavy naphtha (180-350°F), and middle-distillate (350-650°F$^+$). Light naphtha goes to gasoline blending and the middle distillate goes to diesel fuel.

Heavy naphtha is catalytically reformed at high severity (103 RON Clear). The high octane reformate is blended to gasoline. Hydrogen produced in catalytic reforming is compressed and fed to the hydrotreating reactor to reduce the net hydrogen requirements.

Process and economic data on upgrading H-Coal® and CTSL products are based on coal liquid product upgrading work done under DOE contract by Chevron[7] and UOP[8]. Chevron's hydrotreating work has shown, on a variety of coal derived liquids (H-Coal®, SRC, EDS and ITSL), that high severity hydrotreating (LHSV less than 1.0 and 775°F) can produce a middle distillate product with a cetane number greater than 40, and an acceptable quality feedstock for catalytic reforming over bi-metallic reforming catalysts. Chevron is currently extending their data base to include upgrading products from HRI's CTSL Process[9]. Both UOP and Chevron have shown the excellent reformability of coal derived naphthas, with high yields of high octane (103 RON) C_5^+ reformate, with low gas and high hydrogen yields. Product upgrading for this conceptual design is based on HRI's interpretation of this data. Following are the feedstock qualities to product upgrading.

°API	25.4
Carbon	88.05
Hydrogen	11.56
Nitrogen	0.38
Sulfur	0.01
H/C Atomic Ratio	1.58
Kw	10.80
<u>Composition, W %</u>	
C_5-180°F	4.0
180-350°F	14.9
350°F$^+$	81.1

Material balance and utility summaries are provided in Tables 37 and 38.

TABLE 37

AREA 600 - PRODUCT UPGRADING

MATERIAL BALANCE SUMMARY

FEEDS, LB/HR	
Coal Liquids	494,715
Net Make-up Hydrogen*	10,338
TOTAL	505,053
PRODUCTS, LB/HR	
Purge Gases	8,379
Light Naphtha plus Reformate	139,574
Diesel	357,100
TOTAL	505,053

*Net hydrogen make-up. Total hydrogen consumption less hydrogen produced in catalytic reforming.

TABLE 38

AREA 600 - PRODUCT UPGRADING

UTILITIES SUMMARY

Electric Power, Kw	14,915
Cooling Water, Gpm	4,151
Net Steam Consumed	
75 Psig, Lb/Hr	38,589
Process Fuel, MMBtu/Hr	282

AREA 700 - UTILITIES

All of the utility requirements for the liquefaction plant are generated on site, with the exception of purchased electric power. High pressure steam is generated in boilers fired by gases produced in the process. Facilities for electric power distribution are provided. Some major electric power requirements (compressors, etc.) are provided by steam turbines. All water facilities (boiler feedwater, cooling water, process water and potable water) are also provided. Process fuel is internally derived.

Utility usages by plant section are summarized in Table 39. Steam balances are shown in Tables 40.

TABLE 39

UTILITIES

AREA	Power, Kw Electric	Power, Kw Turbine	Cooling Water, Gpm	Process Fuel, MMBtu/Hr	Raw Water, Gpm
100 - Coal Preparation	7,980		132	266	
200 - Liquefaction	63,811		4,520	731	5
300 - Partial Oxidation	12,852		29,847	28	337
400 - Oxygen Plant		65,937	29,299		
500 - Product Treating	2,398		4,830	4	
600 - Product Upgrading	14,915		4,151	282	
700 - Utilities	2,535			160	3,015
800 - Tankage	1,179		24		
TOTAL	105,670	65,937	72,803	1,471	3,357

TABLE 40

STEAM BALANCE

	PSIG 600	PSIG 200	PSIG 75
Net Steam Consumption, Lb/Hr			
AREA			
300 - Partial Oxidation	84,920		
500 - Product Treating	68,511		110,663
600 - Product Upgrading			38,589
700 - Utilities*		10,095	461,348
TOTAL	153,431	10,095	610,600
Net Steam Production, Lb/Hr			
AREA			
200 - Liquefaction	60,767	10,095	
300 - Partial Oxidation			600,505
700 - Utilities	92,664		10,095
TOTAL	153,431	10,095	610,600

*Steam consumed in turbine drivers.

AREA 800 - TANKAGE

Intermediate and final product tankage for plant products and by-products is provided for thirty days. Also included are associated transfer pumps, railcar loading and railcar unloading facilities.

AREA 900 - GENERAL OFFSITES

Included in the plant offsites are electrical systems, interconnecting piping, site preparation, perimeter fencing, roads, parking areas, maintenance equipment, laboratory equipment, mobile equipment, fire protection equipment, buildings, furnishings, communication systems and railroad facilities.

SECTION 5

CAPITAL COST ESTIMATES AND ECONOMIC EVALUATION

CAPITAL COST ESTIMATES AND ECONOMIC EVALUATION

CAPITAL INVESTMENT BASIS

The capital investment estimates are based on the following assumptions:

Plant Size

8,400 TPSD of dry coal to liquefaction. This plant size is based on design studies done by Bechtel for the Breckinridge Project. These studies determined the minimum economic plant size for a commercial H-Coal® liquefaction plant. The minimum plant contains five parallel reactor trains, each processing 1,680 TPSD of dry coal. The capacity per train is set by providing the maximum size reactor (diameter) that can be shop fabricated using conventional techniques.

Capital Cost Basis

1984 dollars at a U. S. Gulf Coast location. This basis was selected to be consistent with prior studies on Illinois and Wyodak coals.

Estimate Basis

Cost estimates are detailed for the various functional areas of the plant. Area 200 - Liquefaction is based on the preliminary designs done by HRI. Costs for major equipment are estimated based on vendor input or recent quotations for similar equipment (see Table 41). Commodity materials and labor are factored based on statistical techniques which HRI has developed for the H-Coal® and H-Oil® Processes. For areas other than liquefaction, curve-type cost estimates are developed based on information contained in HRI's Process Planning Model.

Total Erected Cost

The total erected cost estimated is the sum of direct material and labor, indirect costs and project contingency. Indirect cost is estimated to be 40% of the total direct cost and includes field supervision, equipment and tools, sales tax, and engineering and home office fees. Project contingency is used to allow for the cost of additional equipment that would be specified in a more detailed design of a definitive project at a specific site. The project contingency used is 25%. The total erected costs for the liquefaction section and total plant are detailed in Tables 42. The total erected cost is 2,109 MM$.

ECONOMIC EVALUATION

Product (gasoline and diesel) costs are calculated using a proprietary program developed by HRI for discounted cash flow (DCF) economic calculations. The product costs are based on an economic basis which closely corresponds to those used for major projects proposed for consideration by the U. S. Synthetic Fuels Corporation (SFC). Although this economic basis is not completely valid with the current day situation, it was selected to be consistent with the prior studies on Illinois coal. The economic basis is presented in Table 43 and some highlights are summarized below:

- Debt/Equity Ratio - 75/25 ratio, based on maximum loan guarantees available from SFC.
- Depreciation Method - Accelerated Cost Recovery System (ACRS) based on legislation enacted by Congress in 1981. According to the guidelines of ACRS, a plant of this type can be depreciated over a five year period by the percentages shown below:

Year	% of Investment Depreciated
0	20
1	32
2	24
3	16
4	8

- Investment Tax Credit - 10% taken in the year of investment.

- Working Capital - Estimated at 6% of total erected cost and covers operator training, equipment checkout, major changes in plant equipment, extra maintenance and inefficient use of fuel and other materials that occur during plant start up.

- Inflation - 5% per year on all costs and revenues.

- Production Schedule - Accounts for reduced plant service factor over the initial years of plant operations. It is assumed that the plant operates at 50% of its stated capacity in the first year of operation, 75% in the second and 100% thereafter.

- Interest on Debt - 10%

- DCF Return on Equity - 15%

Based on this assumed economic basis, the equivalent capital charge factor (equivalent to that used in constant dollar economics) is 11% of the total erected cost per year. The capital charge factor is used to calculate a first-year product selling price.

Other operating costs are calculated as follows:

- Coal - $30/Ton Dry Coal

- Electric Power - $0.035/Kwh

- Natural Gas - $2.50/MMBtu

- Raw Water - $0.50/1000 Gallons

- Catalysts and Chemicals (Per Individual Unit Requirements) - Catalyst used in liquefaction is estimated to have a net cost of $2.70/lb after resale for metals reclamation. This is substantially higher than current costs for similar catalysts based on a current highly competitive catalyst market.

- Ash Disposal - $2/Ton

- Labor, Supervision and Overhead - Estimated to be 3% of the total erected cost per year.

- <u>Maintenance</u> - Also estimated at a percentage of the total erected cost. Percentages vary for plant areas as shown below:

Area			Maintenance, % of Total Erected Cost
100	-	Coal Preparation	4.0
200	-	Liquefaction	4.0
300	-	Partial Oxidation	4.0
400	-	Oxygen Plant	2.0
500	-	Product Treating	2.0
600	-	Product Upgrading	2.0
700	-	Utilities	2.0
800	-	Tankage	2.0
900	-	General Offsites	2.0

By-product revenues are included for LPG ($20/Bbl) sulfur ($60/LT) and ammonia ($150/T).

The product costs are tabulated in Table 44. The calculated first-year product cost is $38.35/Bbl.

LIQUEFACTION PLANT INVESTMENT DETAILS

	MAJOR EQUIPMENT COSTS, M$
SECTIONS 100-500 (REACTION)	
J - Pumps	16,725
K - Reactors	46,760
L - Heaters	8,233
M - Exchangers	11,629
Q - Drums	18,608
R - Compressors	23,224
S - Special	12,390
Sub-Total	137,569
SECTIONS 150-550 (CATALYST HANDLING)	
J - Pumps	1,052
M - Exchangers	152
Q - Drums	4,288
S - Special	305
Sub-Total	5,797
SECTION 500 (RECYCLE SLURRY PREPARATION)	
J - Pumps	672
Q - Drums	57
S - Special	13,461
Sub-Total	14,190
SECTION 600 (PRODUCT FRACTIONATION)	
J - Pumps	449
L - Heaters	2,051
M - Exchangers	2,261
N - Towers	934
Q - Drums	394
R - Compressors	467
S - Special	279
Sub-Total	6,835
TOTAL	164,391
TOTAL PLANT INVESTMENT	
Direct Material, MM$	316
Direct Labor, MM$	91
Total Direct Cost, MM$	407
Indirect Cost, MM$	88
Contingency (25%), MM$	124
TOTAL PLANT INVESTMENT, MM$	619

TOTAL PLANT INVESTMENT SUMMARY

AREA	INVESTMENT, MM$*
100 - Coal Preparation	176
200 - Liquefaction	619
300 - Partial Oxidation	283
400 - Oxygen Plant	130
500 - Product Treating	77
600 - Product Upgrading	157
700 - Utilities	109
800 - Tankage	81
900 - General Offsites	477
TOTAL (TEC)	2,109

*1984 U.S. dollars at a U.S. Gulf Coast location.

ECONOMIC BASIS

Project Life, Years		25
Tax Rate, %		50
Debt/Equity Ratio		75/25
Depreciation Method		ACRS
Depreciation, Years		5
Investment Tax Credit, %		10
Working Capital, % of TEC		6
Start-up Expense, % of TEC		6
Inflation Rate, %/Year		5
Capital Expenditure Profile		
% of Investment in Year -	5	5
	4	15
	3	30
	2	35
	1	15
Production Schedule		
% of Capacity in Year -	0	50
	1	75
	2	100
	3, etc.	100
Service Factor, %		90
Interest Rate on Debt, %		10
DCF Return on Equity, %		15

PRODUCT COST

OPERATING COSTS, MM$/YR	
Coal @ $30/Ton	126.8
Electric Power @ $0.035/Kwh	29.3
Raw Water @ $0.50/MGal	0.8
Catalyst and Chemicals	28.8
Ash Disposal @ $2/Ton	1.0
Labor, Supervision and Overhead	63.3
Maintenance	63.7
Capital Charges(1)	232.0
TOTAL	545.7
By-Product Revenues, MM$/Yr	
LPG @ $20/Bbl	1.9
Sulfur @ $60/LT	5.8
Ammonia @ $150/T	7.1
TOTAL	14.8
Product Cost(2)	
MM$/Yr	530.9
$/Bbl	38.35

(1)Equivalent capital charge factor is 11% of total erected cost per year.

(2)First year product cost.

SECTION 6

OVERALL ASSESSMENT OF ECONOMICS

OVERALL ASSESSMENT OF ECONOMICS

It is important to analyze the results of this study to identify the major cost areas in coal liquefaction, their impact on product costs and opportunities for improvement in overall economics. This section will address these topics, while specific recommendations to better utilize some of the unique advantages of the CTSL technology, to further improve economics, will be detailed later.

ANALYSIS OF PRODUCT COSTS

Product costs are analyzed in two different ways to provide a better understanding of the major cost areas in direct coal liquefaction. The first method details the product costs according to the major cost items. Included in the breakdown of cost items are coal, utilities, catalyst and chemicals, labor, supervision and overhead, maintenance, capital charges and by-product revenues. The second method separates the product cost into the respective process areas of the plant including liquefaction, hydrogen production, product upgrading, product treating and offsites.

The breakdown of product cost by cost item is shown in the top part of Table 45. The largest contribution is capital charges at 44% of the total product cost. The next largest contribution is the cost of coal at 24% of the total product cost. Labor, supervision and overhead, and maintenance are both roughly equivalent at 12% of the total product cost. The cost of utilities and catalyst and chemicals are also roughly equivalent at 6% and 5%. By-product revenues reduce product cost by only 3%.

Product costs separated into the major process areas are shown in the bottom part of Table 45. Liquefaction is the largest cost process area at 49% of the total. Hydrogen production and offsites each contribute about 22% of the total product cost. Product upgrading and product treating represent relatively small cost contributions at 7% and 3%, respectively.

CATALYST COST

In the economics presented in Table 45, the catalyst usage amounted to three pounds per ton of dry coal fed to liquefaction, estimated using the rate of catalyst deactivation indicated in a kinetic model based on CTSL bench unit operations. Recent laboratory work, Run 227-34(I-15), successfully demonstrated the feasibility of cascading catalyst from the first stage to the second reactor stage, while maintaining overall process performance.[1] Catalyst cascading was projected to reduce consumption by 20%. This would reduce catalyst and chemicals cost from $2.08/Bbl to $1.76/Bbl and the overall product cost from $38.35/Bbl to $38.03/Bbl.

It should be noted that catalyst usage reported for the light recycle CTSL previous case was a projected number of one pound per ton, to be comparable with H-Coal. The actual usage demonstrated in the previous case was 3 Lbs/Ton. A usage of 2.19 Lbs/Ton, without cascading, would apply for the light recycle case, if the catalyst usage for the two cases were taken to be in inverse ratio of the catalyst ages for the CTSL operations upon which the process designs were based. Table 46 shows the revised product cost reflecting both catalyst cascading for the heavy recycle case, and the higher catalyst consumption for the light recycle case, without cascading.

Comparison With Previous Illinois Coal Studies

Table 47 presents economics of the heavy recycle CTSL Process using Illinois No. 6 coal, compared to light recycle CTSL and to one-stage H-Coal®. Presented also are hydrogen production alternatives of partial oxidation versus steam reforming for the two CTSL base cases, namely heavy and light recycle[10].

Heavy recycle CTSL is seen resulting in over 5% product cost savings compared to light recycle CTSL and almost 17% compared to one stage H-Coal®. The savings occur despite the higher required capital investment for heavy recycle, in part because of the higher liquid product yield from coal, namely 3.28 versus 3.18 barrels per ton of total coal feed and, in part, because of the lower investment per unit of product, namely 50,280 $/BPSD versus 54,900 $/BPSD.

Heavy recycle appears more economical than light recycle regardless of hydrogen production method, as discussed below.

Impact of Coal Cleaning Efficiency

Coal supplied to liquefaction is deep-cleaned using heavy media washing to reduce ash content to the required level. Based on experimental work done by BCR National Laboratory on this Illinois No. 6 coal and on a published report by the Electric Power Research Institute (EPRI)[11], an overall coal cleaning efficiency of 70% was assumed for this study. While the reject stream is utilized in hydrogen manufacture, a 70% efficiency means that for every ton of coal liquefied 1.52 tons are furnished to coal cleaning. In addition hydrogen manufacture by partial oxidation of a coal water slurry is recognized to be a highly capital-intensive approach.

Table 48 presents the economic incentive for increasing coal cleaning efficiency. A considerable decrease is shown both for the total capital investment and the overall product cost. Thus, if efficiency is increased from 70% to 80%, the product cost decreases from $38.35/Bbl to $36.69/Bbl. Figure 4 shows the product cost as a function of coal cleaning efficiency.

Figure 4 assumed that hydrogen shortfall from partial oxidation of the reject stream from deep-cleaning coal at higher efficiency would be supplied by steam reforming purchased natural gas. This hydrogen could be supplied by partial oxidation of additional coal, but as seen from the middle two columns of Table 48, this is a more expensive alternate using $30/Ton coal and natural gas at $2.50/MMBtu. In other words, the justification for coal partial oxidation is that the coal stream is already present as a waste product. If the stream is lost by a more efficient coal cleaning operation, partial oxidation of additional coal becomes cost ineffective.

Figure 5 is a breakeven cost comparison of hydrogen production alternatives at 80% coal cleaning efficiency. Thus, if coal is available at $30/Ton, steam reforming would be preferred over partial oxidation for any gas price below $3.77/MMBtu.

Impact of Purchasing Deep-Cleaned Coal

The basic assumption made in this study was that low ash-level coal, required for liquefaction, would not be commercially available and, therefore, the deep cleaning operation would be performed on site. Credit was takenfor the high-ash rejection stream from the deep-cleaning operation, in supplying a part of the make-up hydrogen via partial oxidation.

Assuming that deep-cleaned coal is available on the open market, the economic impact of purchasing this deep-cleaned Illinois No. 6 coal was investigated in two alternatives:

1. Generating hydrogen by steam reforming purchased natural gas, or

2. Generating hydrogen by partial oxidation of deep-cleaned coal.

Table 49 shows the economics of the steam reforming alternative compared to the base case. It is seen that this alternative will lower the total plant investment. This is due to the high cost of both the partial oxidation and the oxygen plant, as well as the cost of the coal preparation plant. Consequently, product cost is lowered by using steam reforming.

One of the major unknowns in analyzing the economics of these alternatives is the cost of deep-cleaned coal. The delivered price of Illinois No. 6 coal containing 11.7% ash has been assumed in the base case to be $30/Ton. Equating this price on an equal-BTU basis would result in a deep-cleaned coal price of $31.42/Ton. However, rough studies using published data on deep-cleaning economics would indicate a cost of around $35.50/Ton. The comparative economics of the two alternatives were evaluated with varying coal and natural gas prices. Figure 6 shows using deep-cleaned coal in hydrogen manufacture and in liquefaction to be an economic standoff with providing for deep-cleaning at the plant site.

Figure 7 shows that eliminating coal washing at the plant site and purchasing deep-cleaned coal and using steam reforming for hydrogen manufacture appears to offer significant cost reduction over the base case, with fuel cost at $2.50/MMBtu and with any deep-cleaned coal cost below $43/Ton. Alternatively, with deep-cleaned coal at $35/Ton this option will offer an economic advantage over the base case for any fuel cost below $3.17/MMBtu.

Hydrogen Production Alternatives

In the previous study on light recycle CTSL processing of Illinois coal, steam reforming compared favorably with partial oxidation in many pricing situations. Using a coal cleaning efficiency of 70% for the heavy recycle operation, however, such is not the case, as seen in Table 47. Figure 8 shows the effect of natural gas price on product cost. The breakeven natural gas cost is around \$2.05/MMBtu, compared to \$2.77/MMBtu in light recycle operation. This is primarily due to the particular balance achieved between hydrogen requirement and hydrogen availability from the coal cleaning reject stream. As noted above, however, steam reforming is the less costly alternative when coal cleaning efficiency is increased.

Coal Cost

The base coal cost is \$30/Ton. Figure 9 shows that regardless of coal price, heavy recycle CTSL is less costly than light recycle CTSL (using either partial oxidation or steam reforming)and H-Coal®. Since coal to the complex remains the same in heavy recycle, hydrogen production method does not enter into this analysis.

Comparison With Wyoming Coal

A previous study had compared CTSL economics using Illinois coal and light recycle with Wyoming coal[12]. Table 49 updates the previous study to include Illinois coal heavy recycle CTSL operation. Comparing heavy recycle Illinois coal CTSL with Wyodak coal CTSL, it is seen that using steam reforming to generate hydrogen Wyoming coal has very slightly better economics (\$39.30/Bbl versus \$39.45/Bbl). Using partial oxidation to generate hydrogen Illinois No. 6 coal, with either light or heavy recycle, is less costly than Wyodak coal.

Figure 10 presents the sensitivity of product cost to natural gas price for Wyodak coal and for Illinois coal with heavy recycle. The Illinois coal/natural gas price situation has already been discussed above and is repeated here only for comparison with Wyodak coal. Figure 10 shows that for any gas price Illinois coal is preferred over Wyodak coal when hydrogen is generated by partial oxidation. For hydrogen via steam reforming the two coals are equivalent.

The above economic comparison of Wyodak versus Illinois coals assumed $10/Ton for the former and $30/Ton for the latter. Figure 11 presents a breakeven analysis for the coal price of Wyodak coal versus Illinois coal with heavy recycle CTSL processing. Two lines are shown for the two hydrogen production alternatives. For a given delivered Illinois coal price, a Wyodak coal price above the respective hydrogen-producing-method line would favor Illinois coal while below the line would favor use of Wyodak coal. Of course, Figure 11 could also be used the other way, entering a given Wyodak coal price. In general, with steam reforming for hydrogen manufacture, Illinois coal is to be preferred if its cost differential over that of Wyodak coal is less thant $21-24/Ton. With partial oxidation for hydrogen manufacture, this actual cost differential is $28-33/Ton.

In summary, Illinois coal at $30/Ton, using heavy recycle, is preferred, to Wyodak coat at $10/Ton, for any of the hydrogen manufacturing methods considered. For the light recycle operation the relative economics depend on the hydrogen production method.

BREAKDOWN OF PRODUCT COST

	CTSL Heavy Recycle $/Bbl	CTSL Heavy Recycle %	CTSL Light Recycle $/Bbl	CTSL Light Recycle %	H-Coal® $/Bbl	H-Coal® %
BY COST ITEM						
Coal	9.16	24	9.43	23	10.97	24
Utilities*	2.25	6	2.29	6	2.83	6
Catalyst and Chemicals	2.08	5	1.21	3	1.49	3
Labor, Supervision and Overhead	4.57	12	4.93	12	6.04	13
Maintenance	4.60	12	5.07	13	6.18	14
Capital Charges	16.76	44	18.64	46	22.94	50
By-Product Revenue	(1.07)	(3)	(1.05)	(3)	(4.51)	(10)
TOTAL	38.35	100	40.52	100	45.94	100
BY PROCESS AREA**						
Liquefaction	18.84	49	19.69	49	25.43	56
Hydrogen Production	8.66	22	7.41	18	7.30	15
Product Upgrading	2.59	7	2.63	7	2.96	6
Product Treating	1.00	3	1.58	4	2.38	5
Offsites	8.33	22	10.26	25	12.38	28
By-product Revenue	(1.07)	(3)	(1.05)	(3)	(4.51)	(10)
TOTAL	38.35	100	40.52	100	45.94	100

*Includes electric power, raw water and ash disposal.
**Each process area includes the cost of feedstocks, utilities, catalyst, labor, maintenance and capital.

REVISED BREAKDOWN OF PRODUCT COST

	CTSL					
	Heavy Recycle		Light Recycle		H-Coal®	
	$/Bbl	%	$/Bbl	%	$/Bbl	%
BY COST ITEM						
Coal	9.16	24	9.43	23	10.97	24
Utilities*	2.25	6	2.29	6	2.83	6
Catalyst and Chemicals**	1.76	5	2.07	5	1.49	3
Labor, Supervision and Overhead	4.57	12	4.93	12	6.04	13
Maintenance	4.60	12	5.07	12	6.18	14
Capital Charges	16.76	44	18.64	45	22.94	50
By-Product Revenue	(1.07)	(3)	(1.05)	(3)	(4.51)	(10)
TOTAL	38.03	100	41.38	100	45.94	100
BY PROCESS AREA* **						
Liquefaction	18.52	49	20.55	50	25.43	56
Hydrogen Production	8.66	22	7.41	18	7.30	15
Product Upgrading	2.59	7	2.63	6	2.96	6
Product Treating	1.00	3	1.58	4	2.38	5
Offsites	8.33	22	10.26	25	12.38	28
By-product Revenue	(1.07)	(3)	(1.05)	(3)	(4.51)	(10)
TOTAL	38.03	100	41.38	100	45.94	100

*Includes electric power, raw water and ash disposal.

**Catalyst consumption in heavy recycle reflects the effect of catalyst cascading. Catalyst consumption in light recycle adjusted to reflect equivalent catalyst ages for the two CTSL cases, without cascading.

***Each process area includes the cost of feedstocks, utilities, catalyst, labor, maintenance and capital.

SUMMARY OF RESULTS OF ECONOMIC EVALUATIONS

PROCESS	CTSL	CTSL	CTSL	CTSL	H-COAL®
RECYCLE	Heavy	Heavy	Light	Light	-
HYDROGEN MANUFACTURE(1)	POX	SR	POX	SR	POX
COAL FEED, TPSD					
To Liquefaction	8,400	8,400	9,428	9,428	9,428
To Partial Oxidation	4,405	-	1,552	-	-
Power Generation, etc.	-	4,405	-	-	-
TOTAL	12,805	12,805	10,980	9,428	9,428
PURCHASED UTILITIES					
Power, MW	106	-	90	85	82
Natural Gas, MMSCFD	-	102	-	39	-
LIQUID PRODUCTS, BPSD					
Gasoline	13,170	13,170	10,967	10,967	9,700
Diesel	28,778	28,778	23,967	23,967	16,078
TOTAL	41,948	41,948	34,934	34,934	25,778
Total Plant Investment, \$MM(2)	2,109	1,882	1,918	1,765	1,738
\$/BPSD	50,280	44,870	54,900	50,520	67,420
Product Cost, \$/Bbl(3)	38.35	39.45	40.52	40.23	45.94

(1) POX = Partial Oxidation, SR = Steam Reforming.
(2) 1984 total erected cost at U.S. Gulf Coast location includes 25% project contingency.
(3) Based on coal at \$30/Ton and natural gas at \$2.50/MMBtu. Other bases listed in Section 5.

EFFECT OF COAL CLEANING EFFICIENCY ON PROJECT ECONOMICS

ASSUMED COAL CLEANING EFFICIENCY, %	70	80	80	90
SUPPLEMENTAL H_2 MANUFACTURE[1]	-	SR	POX	SR
TOTAL COAL REQUIRED, TPSD	12,805	11,205	12,866	9,960
PURCHASED UTILITIES				
Power, MW	106	92	80	77
Natural Gas, MMSCFD	-	41	-	75
Total Plant Investment, $MM[2]	2,109	1,891	2,112	1,723
PRODUCT COST, $/BBL[3]				
Coal	9.16	8.01	9.20	7.12
Utilities	2.25	4.42	1.75	6.11
Catalyst and Chemicals	2.08	2.05	2.08	2.03
Labor, Supervision and Overhead	4.57	4.10	4.58	3.73
Maintenance	4.60	4.10	4.61	3.72
Capital Charges	16.76	15.03	16.78	13.69
By-Product Revenue	(1.07)	(1.02)	(1.07)	(0.99)
	38.35	36.69	37.93	35.41

(1)POX = Partial Oxidation, SR = Steam Reforming.

(2)1984 total erected cost at U.S. Gulf Coast location.

(3)Based on $30/Ton for coal, natural gas at $2.50/MMBtu. Other bases listed in Section 5.

COMPARISON OF ILLINOIS WITH WYODAK COAL

COAL FEED	ILLINOIS	ILLINOIS	ILLINOIS	ILLINOIS	WYODAK	WYODAK
Recycle	Heavy	Heavy	Light	Light		
Hydrogen Manufacture(1)	POX	SR	POX	SR	POX	SR
COAL FEED, TPSD						
To Liquefaction	8,400	8,400	9,428	9,428	10,000	10,000
To Partial Oxidation	4,405	-	1,552	-	3 650	-
Supplemental To Fuel	-	749	-	-	1,068	-
Excess Coal Cleaning Waste		3,656				
TOTAL	12,805	12,805	10,980	9,428	14,718	10,000
PURCHASED UTILITIES						
Power, MW	106	-	90	85	115	109
Natural Gas, MMSCFD	-	102	-	39	-	92
LIQUID PRODUCTS						
Gasoline	13,170	13,170	10,967	10,967	12,660	12,660
Diesel	28,778	28,778	23,967	23,967	24,160	24,160
Turbine Fuel	-	-	-	-	948	948
TOTAL	41,948	41,948	34,934	34,934	37,768	37,768
Total Plant Investment, $MM(2)	2,109	1,882	1,918	1,765	2,391	1,887
Product Cost, $/Bbl(3)	38.35	39.45(4) 36.99(5)	40.52	40.23	41.55	39.30

(1) POX = Partial Oxidation, SR = Steam Reforming.
(2) 1984 total erected cost at U.S. Gulf Coast location.
(3) Based on $30/Ton for Illinois Coal, $10/Ton for Wyodak Coal and Natural Gas at $2.50/MMBtu. Other bases listed in Section 5.
(4) Coal cleaning waste to waste disposal.
(5) Coal cleaning waste resold at equivalent fuel vlue ($26/Ton).

FIGURE 4

EFFECT OF COAL CLEANING EFFICIENCY ON PRODUCT COST
Heavy Recycle CTSL with Illinois No. 6 Coal

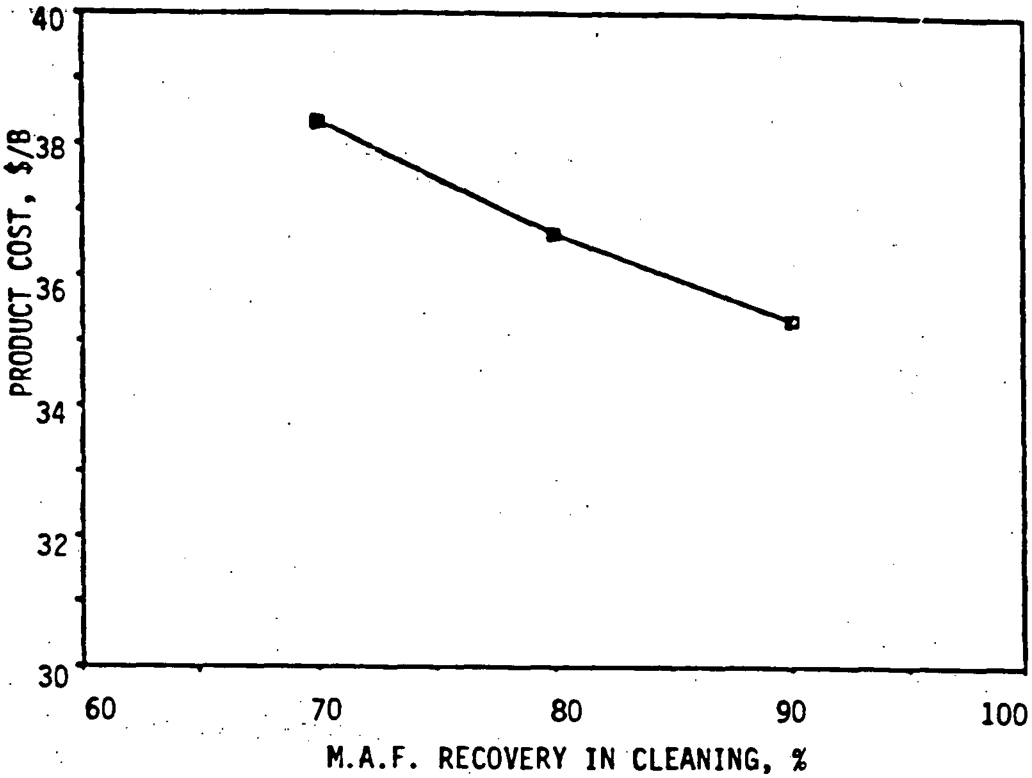


FIGURE 5

BREAKEVEN FEEDSTOCK PRICES, STEAM REFORMING VERSUS PARTIAL OXIDATION, AT 80% COAL CLEANING EFFICIENCY
Heavy Recycle CTSL with Illinois No. 6 Coal

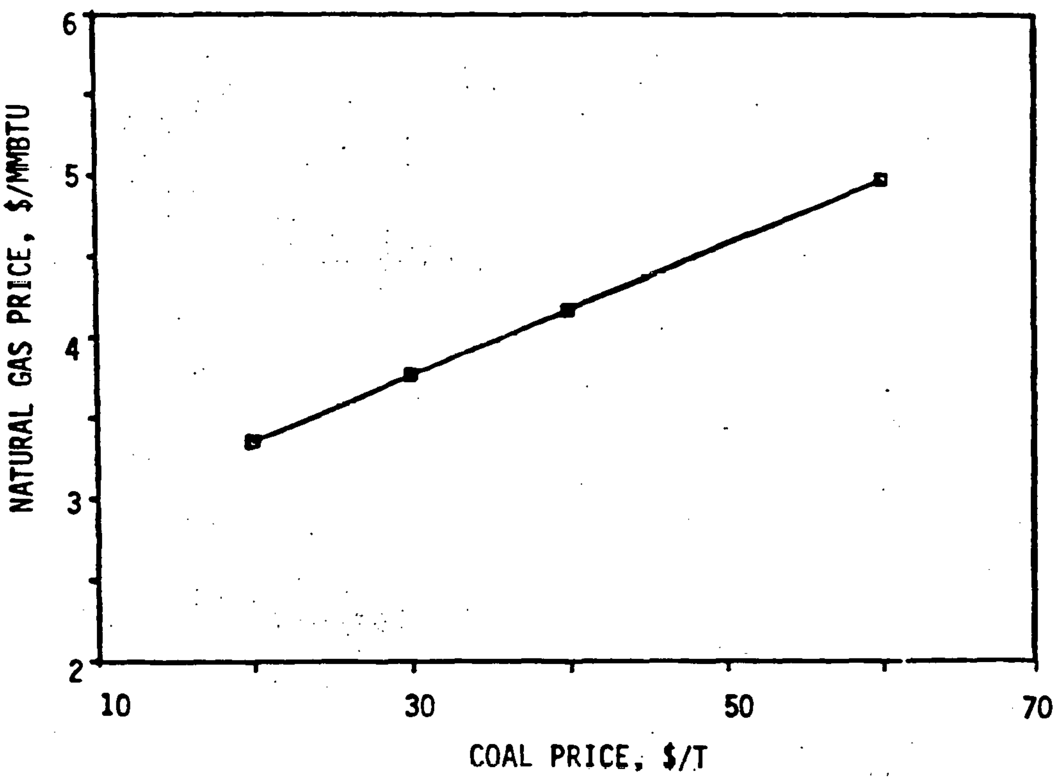


FIGURE 6

EFFECT OF DEEP-CLEANED COAL PRICE ON PRODUCT COST

USING LOW-ASH COAL TO LIQUEFACTION AND POX

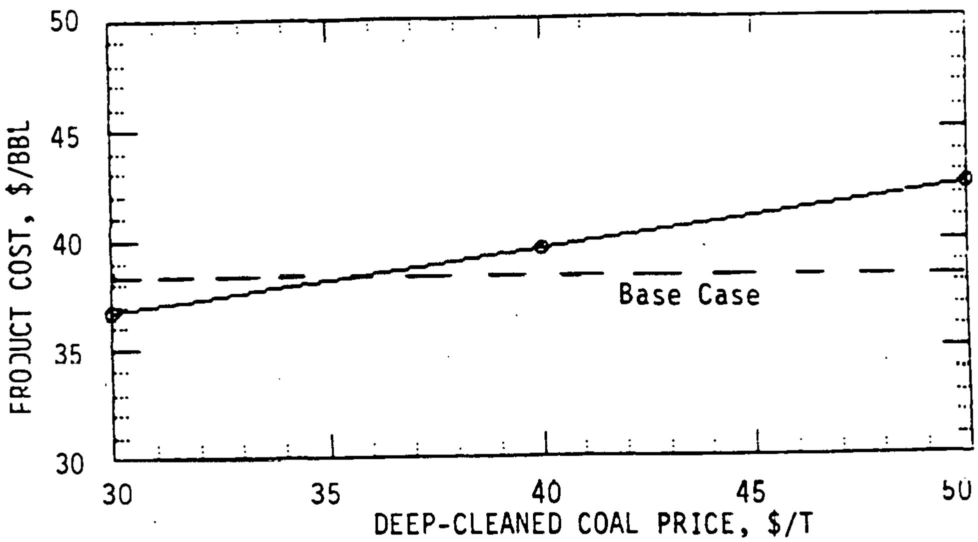


FIGURE 7

USING LOW-ASH COAL TO LIQUEFACTION AND SR FOR H_2 MANUFACTURE

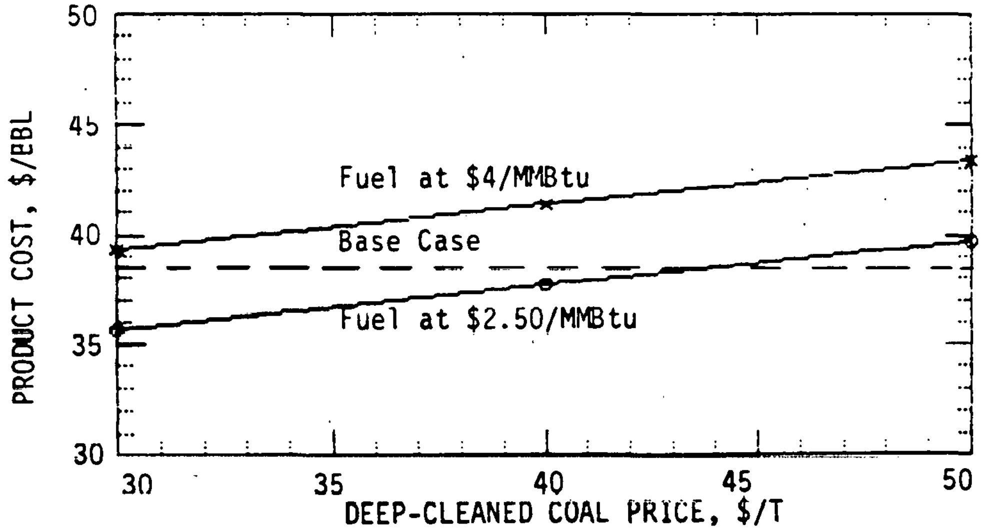


FIGURE 8

EFFECT OF NATURAL GAS PRICE ON PRODUCT COST

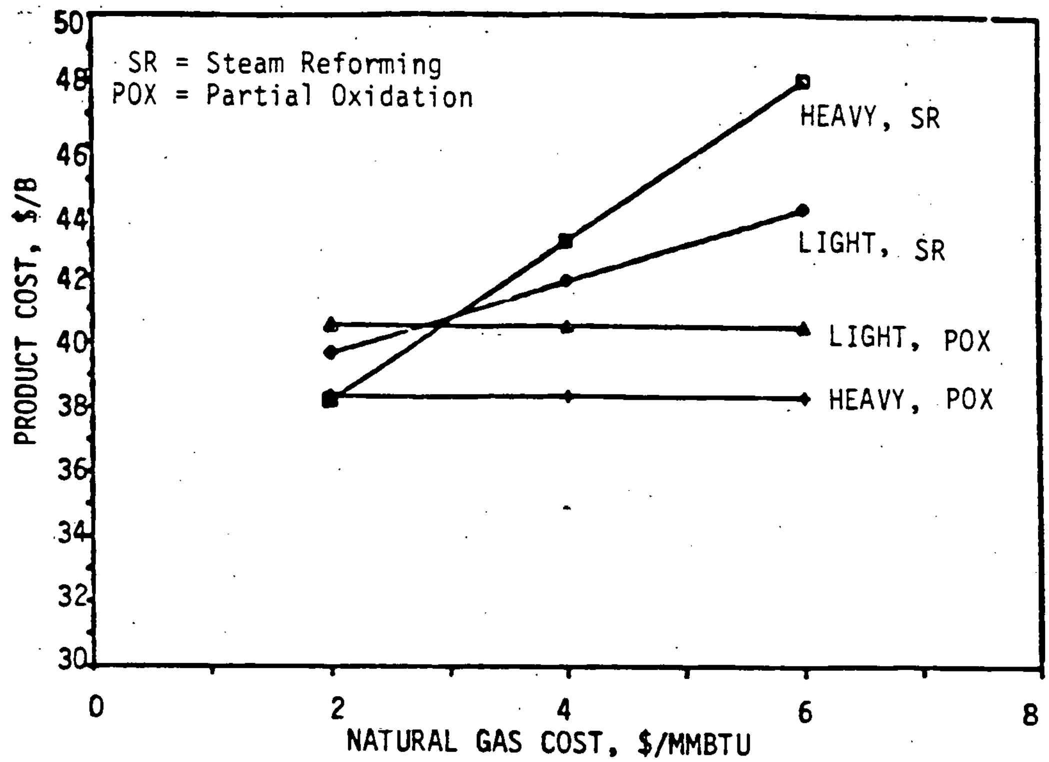
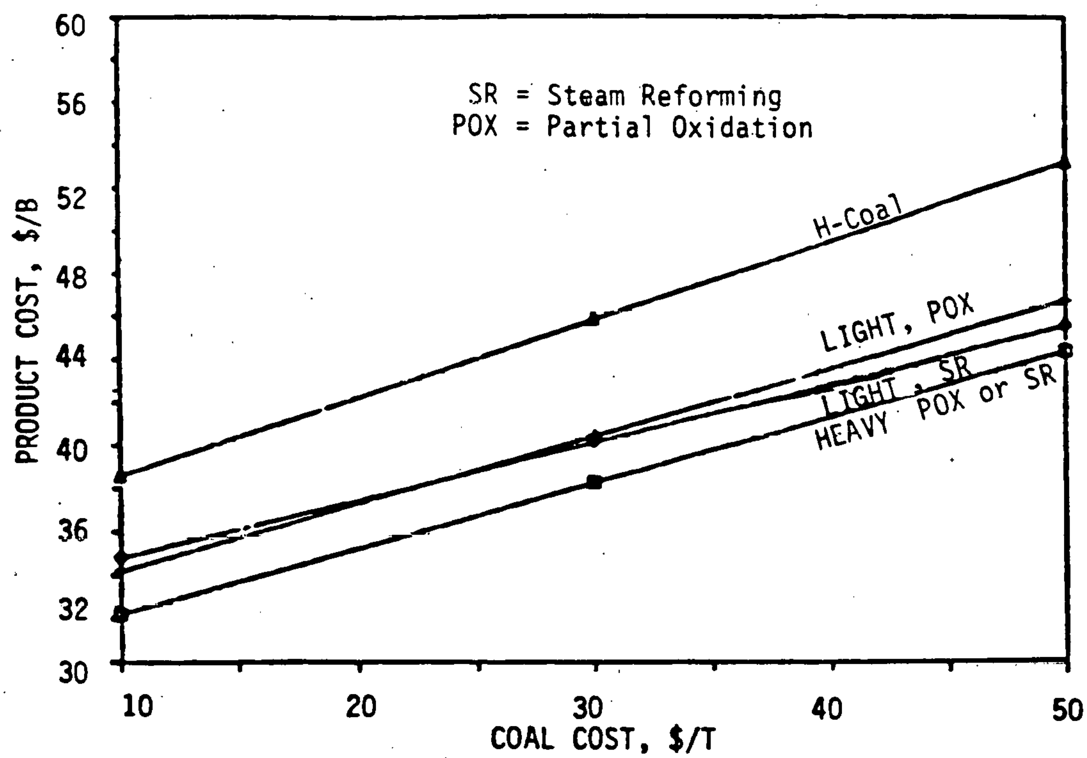


FIGURE 9

EFFECT OF COAL PRICE ON PRODUCT COST



EFFECT OF NATURAL GAS PRICE ON PRODUCT COST

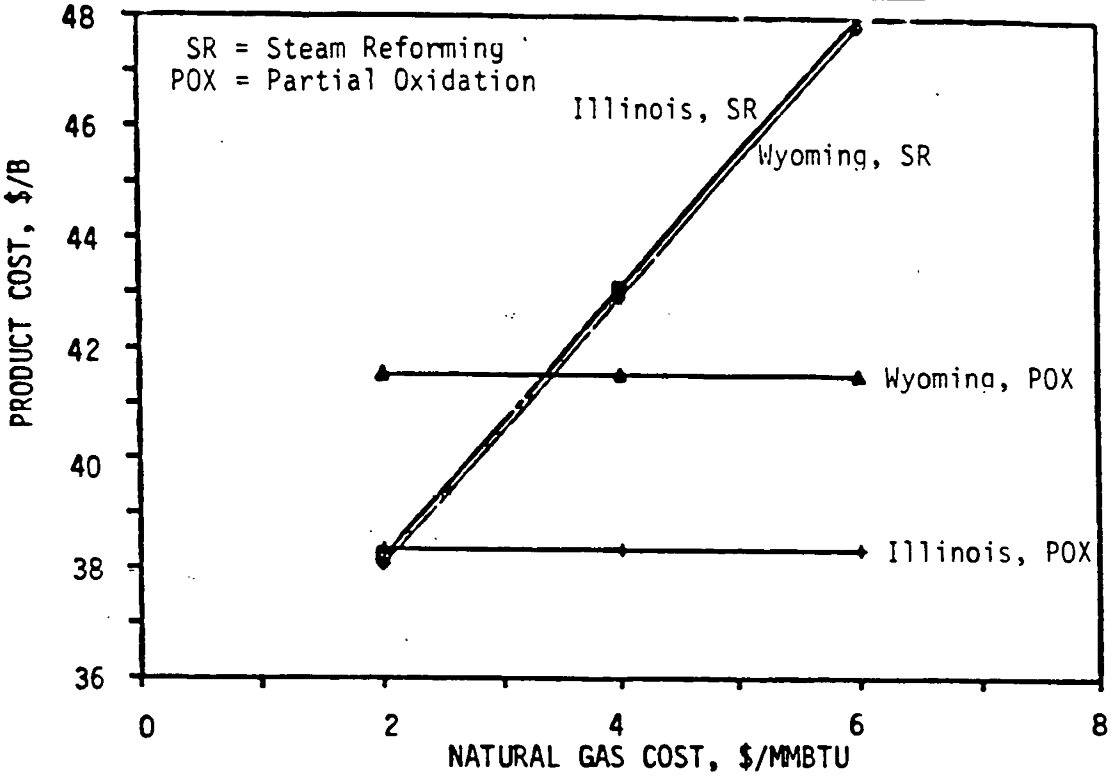
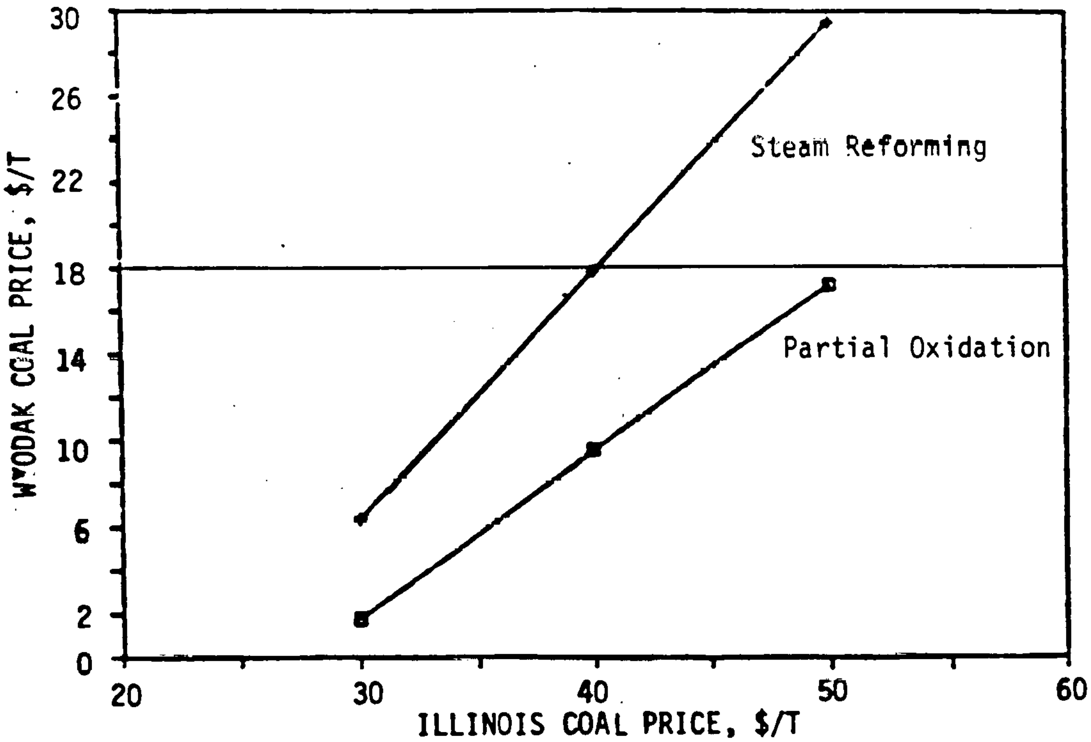


FIGURE 11

BREAKEVEN COAL COST COMPARISON OF ILLINOIS NO. 6 VERSUS WYODAK COALS



SECTION 7

CONCLUSIONS AND RECOMMENDATIONS

CONCLUSIONS AND RECOMMENDATIONS

HRI's Catalytic Two-Stage Liquefaction (CTSL) Process has demonstrated greatly improved process performance compared to the single-stage H-Coal® Process using both Illinois No. 6 and Wyoming sub-bituminous coals. An earlier study comparing light recycle CTSL operation with H-Coal® Process showed a 12% economic advantage for CTSL and identified a number of areas for process and design improvements. These resulted in the heavy recycle CTSL Process with deep-cleaned coal, the economic results of which are contained in this study. These results shows:

- A further 5% economic advantage over light recycle processing and an overall 17% advantage over H-Coal®.
- Product cost for heavy recycle CTSL could be further reduced by 4%, from $38.35/Bbl to $36.69/Bbl if coal-cleaning efficiency is increased from 70% to 80%.
- At increased coal cleaning efficiency, hydrogen manufacture by steam reforming natural gas is preferred over partial oxidation of supplemental coal.
- Catalyst cascading could reduce consumption by 20% and lower product cost from $38.35/Bbl to $38.03/Bbl.
- The price advantage of heavy recycle CTSL processing over light recycle is independent of Illinois No. 6 coal price or natural gas price.
- Partial oxidation is preferred over steam reforming at the assumed 70% coal cleaning efficiency, unless natural gas price falls below $2.05/MMBtu with Illinois No. 6 coal at $30/Ton.
- Illinois coal is a more economically attractive liquefaction feedstock than Wyodak coal: its cost differential over Wyodak coal is less than $28/Ton with partial oxidation to produce hydrogen; the required coal cost differential is lowered to $21/Ton, with steam reforming to produce hydrogen.

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ADDENDUM

CATALYST REQUIREMENTS
FOR
CTSL DEMONSTRATION OPERATION WITH ILLINOIS NO. 6 COAL

The equations for the kinetic model developed for CTSL operations with Illinois No. 6 coal have been used for estimation of the catalyst replacement requirements to obtain the residual oil yield obtained in the design basis operation using the Run 227-47, Period 21, yields. Direct application of the of correlating parameters (which were based on Run 227-30 and Run 227-32 process variable runs) indicated a minimum catalyst requirement of 2.02 Lbs/Ton dry coal (1.52 pounds to the first stage, and 0.50 pounds to the second stage) or 2.17 Lbs/Ton with equal amounts of catalyst to each stage. A comparison of the Run 227-47 results with those projected using the correlating parameters, suggested that catalyst deactivation in that run might have been more severe than reflected by the correlating parameters. With adjustment of the catalyst deactivation parameters of the correlation the estimated catalyst replacement increased to 2.69 Lbs/Ton dry coal (2.00 pounds and 0.69 pounds) or 2.89 Lbs/Ton with equal amounts of catalyst to each stage.

MODEL PARAMETERS AND YIELD CALCULATION METHOD

The calculations were based on the reaction rate equation for the key catalytically dependent reaction in the model, the hydrogenation of the primary low-hydrogen content dissolved residuum component to a secondary hydrogenated residual oil component. In the model, the dissolution of the coal forming the primary residual component, and the cracking of the secondary residual oil component to lighter products were found not to be dependent upon catalyst activity.

In the model, following first order CSTR kinetics based on the weight fraction of the reacting component in the the slurry phase, the key parameters for the primary residual oil component are as follows:

Formation - 50 W % Of Converted Coal

Reaction rate to hydrogenated secondary component follows the equation

K = K1 * exp(-D1 * Age) + K2 * exp(-D2*Age)
K1 = 130 Lbs/Hr/Ft^3 at 750°F
K2 = 11 Lbs/Hr/Ft^3 at 750°F

Activation energy for both rate constant factors - 20 Kcal/mole

D1 = 0.071 * (1.04**(T-750)) D2 = 0.0033 * (1.04**(T-750))

Age - Time of operation with coal feed, Days

This equation rationalizes the relatively high hydrogenation rates, with a relatively moderate decline in the rate constant with increasing catalyst age, in the lower temperature first stage, and the somewhat higher apparent loss of activity in the higher temperature second stage, but with retention of a measure of reaction potential even at the higher catalyst ages. Figure 1 of this addendum summarizes the rate constants against effective catalyst age for the correlated data with a curve showing the fit of the above equations to the experimental results.

The rate constant for the cracking to lighter products of the secondary hydrogenated residual was 15.5 Lbs/Hr/Ft^3 at 750°F, with an activation energy of 16.5 Kcal/mole.

In using the above equations to the system with continuous replacement of the catalyst, the reaction rate constant, Kt, was calculated by the relation

Kt = [K1 * Cat / (Cat + D1)] + [K2 * Cat / (Cat + D2)]

where Cat is the catalyst replacement rate in the same units as the deactivation factors, D1 and D2, in this case fraction replacement of the catalyst inventory per day. The replacement rates in Lbs/Ton dry coal were converted to fractional replacement rates by taking the catalyst inventory to be the same as the demonstration run, 16.5 Lbs/Ft^3 reactor volume.

The calculation of residual oil yields used the conditions that were used during the Run 227-47, Period 21, operation, namely, the same stage temperatures, dry coal feed rate, recycle of pressure filter liquid to the first stage and as buffer flow to the second stage, and residual oil content of the pressure filter liquid. The calculation required setting the slurry phase flow from each stage, which was done by taking the vapor phase flow to be the same as obtained from the second stage hot separator during the demonstration operation.

MATCHING DEMONSTRATION RUN RESULTS WITH CORRELATION PROJECTIONS

The target of the calculations was to match the experimental residual oil yield for the demonstration operation, 2.54 W % of dry coal. Initially, the correlation and calculation method were tested against the results during Run 227-47, Period 21. For a case that assumed a degree of catalyst deactivation that would have occurred had the entire run been at the temperatures holding at that stage of the run, the residual oil yield was calculated as 6.78 W % of dry coal, corresponding to a greater degree of deactivation than indicated experimentally. Projections using the model parameters indicated that the 2.54 W % yield would was to be expected at 18 days of operation with constant stage temperatures maintained as those holding in the twenty-first day.

Such a deviation is to be expected since in Run 227-47 the temperatures of each of the stages had been increased by 25-30°F during the course of the operations so as to sustain the bottom extinction objective of the demonstration operation. Consequently, following the model, catalyst deactivation should have been somewhat less than that projected assuming that final temperature had been maintained from the beginning of the coal operations. Using the relations given above of the deactivation factors, D1 and D2 above, and temperature, it was estimated that degree of catalyst deactivation to be expected at 21 days of the demonstration operation with progressively increasing temperatures, should have been the same as after 13.5 days of operations in which the final temperature had been maintained from the beginning of coal operations.

The discrepancy between the nominal 18 days of deactivation, corresponding to the experimental residual oil yield, and the expected 13.5 equivalent days of deactivation based on the pattern of temperatures earlier in the run, corresponds to an experimental catalyst deactivation somewhat more severe than indicated by the parameters of the model, by a factor of 1.33.

PROJECTIONS WITH CONTINUOUS CATALYST REPLACEMENT

Calculations using the unmodified parameters of the correlation gave the following first and second stage catalyst requirements to obtain the target of 2.54 W % yield of residual oil:

Catalyst Addition, Lb/Ton			Hydrogenation Rate Constants	
Stages			Stages	
First	Second	Total	First	Second
1.080	1.080	2.160	24.96	9.31
1.130	1.000	2.130	25.66	8.73
1.275	0.750	2.025	27.82	7.05
1.518	0.500	2.018	31.22	4.65
1.619	0.400	2.019	32.56	3.78

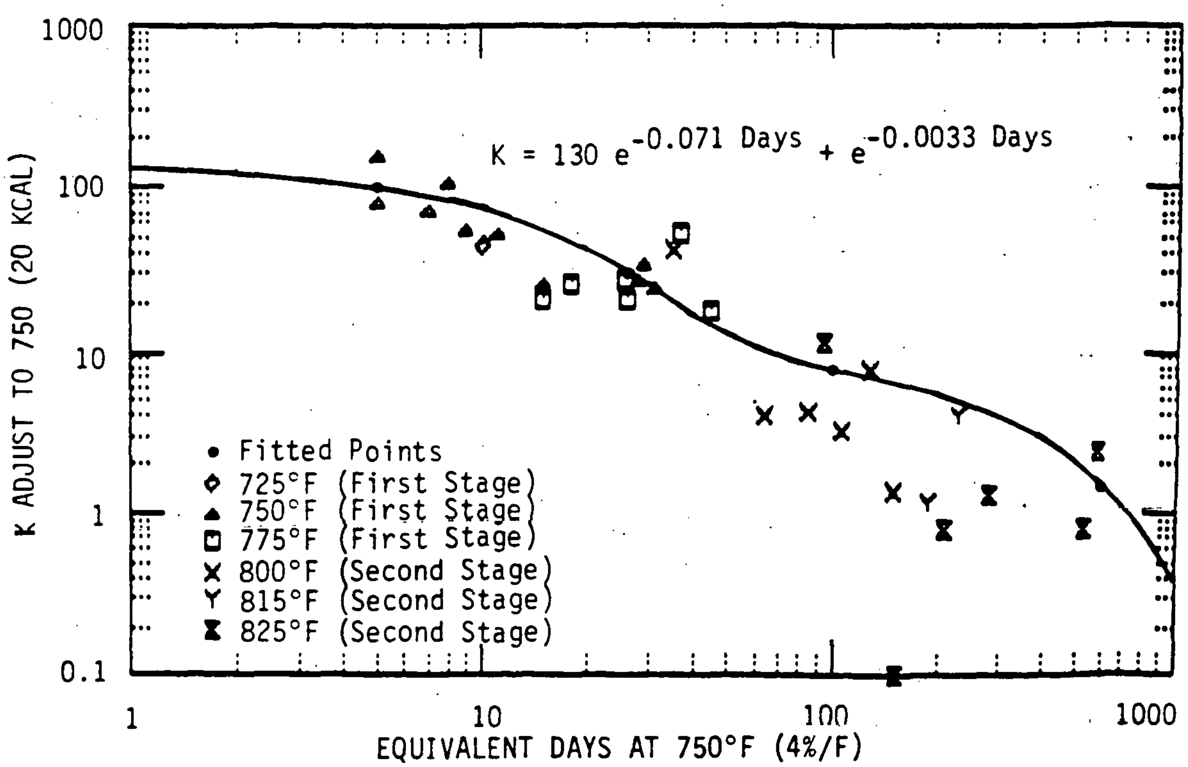
These calculations show a flat minimum in total catalyst requirement at 2.02 Lbs/Ton dry coal, with about one-fourth of the catalyst going to the second stage. With equal amounts of catalyst added to the two stages total catalyst requirement increases by about 7%.

Calculations with the deactivation factors increased by 33% over the original values in the kinetic model correlation gave the following results:

Catalyst Addition, Lb/Ton			Hydrogenation Rate Constants	
Stages			Stages	
First	Second	Total	First	Second
1.444	1.444	2.888	24.94	9.34
1.500	1.347	2.847	25.58	8.81
2.000	0.690	2.690	30.97	4.81
2.250	0.451	2.701	33.44	3.23
2.500	0.250	2.750	35.81	1.83

Again, there is a flat minimum in the total requirement, here at about 2.69 Lbs/Ton, with about one-fourth of the catalyst going to the second stage. With equal amounts of catalyst added to the two stages the total catalyst requirement increases by about 7%.

ADJUSTED HYDROGENATION RATE CONSTANT
VERSUS
EQUIVALENT DAYS OF OPERATION



DOE/PC/80002-T4

LOW-SEVERITY CATALYTIC TWO-STAGE LIQUEFACTION PROCESS:
ILLINOIS COAL CONCEPTUAL COMMERCIAL PLANT DESIGN AND ECONOMICS

DOE