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AFWAL-TR-87-2042
Volume VI

PRODUCTION OF JET FUELS FROM COAL DERIVED LIQUIDS

VOL VI - Preliminary Analysis of Upgrading Alternatives For
The Great Plains Liquid By-Product Streams



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SEPTEMBER 1988

INTERIM REPORT FOR THE PERIOD MARCH 1987 - FEBRUARY 1988

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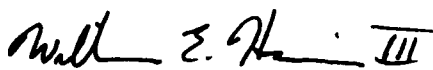
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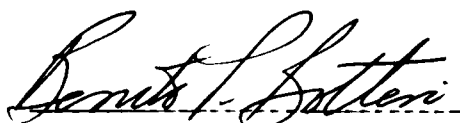
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UNCLASSIFIED
SECURITY CLASSIFICATION OF THIS PAGE

REPORT DOCUMENTATION PAGE				Form Approved OMB No. 0704-0188	
1a. REPORT SECURITY CLASSIFICATION Unclassified			1b. RESTRICTIVE MARKINGS None		
2a. SECURITY CLASSIFICATION AUTHORITY N/A			3. DISTRIBUTION/AVAILABILITY OF REPORT Approved for Public Release; Distribution is Unlimited		
2b. DECLASSIFICATION/DOWNGRADING SCHEDULE N/A					
4. PERFORMING ORGANIZATION REPORT NUMBER(S) N/A			5. MONITORING ORGANIZATION REPORT NUMBER(S) AFWAL-TR-87-2042 VOL VI		
6a. NAME OF PERFORMING ORGANIZATION Amoco Oil Company		6b. OFFICE SYMBOL (If applicable) N/A	7a. NAME OF MONITORING ORGANIZATION Air Force Wright Aeronautical Lab Aero Propulsion Lab (AFWAL/POSF)		
6c. ADDRESS (City, State, and ZIP Code) P. O. Box 400 Naperville, IL 60566			7b. ADDRESS (City, State, and ZIP Code) Wright-Patterson AFB, Ohio 45433-6563		
8a. NAME OF FUNDING/SPONSORING ORGANIZATION		8b. OFFICE SYMBOL (If applicable)	9. PROCUREMENT INSTRUMENT IDENTIFICATION NUMBER FY 1455-86-NO657		
8c. ADDRESS (City, State, and ZIP Code)			10. SOURCE OF FUNDING NUMBERS		
			PROGRAM ELEMENT NO. 63216F	PROJECT NO. 2480	TASK NO. 16
11. TITLE (Include Security Classification) Production of Jet Fuels from Coal Derived Liquids - Vol. VI - Preliminary Analysis of Upgrading Alternatives for the Great Plains Liquids by-Product Streams					
12. PERSONAL AUTHOR(S) B. A. Fleming, J. D. Fox, M. W. Furlong, J. g. Masin, L. P. Sault, D. F. Tatterson					
13a. TYPE OF REPORT Interim		13b. TIME COVERED FROM 3/87 TO 2/88		14. DATE OF REPORT (Year, Month, Day) 1988 September	
15. PAGE COUNT 310					
16. SUPPLEMENTARY NOTATION None					
17. COSATI CODES			18. SUBJECT TERMS (Continue on reverse if necessary and identify by block number) Turbine Fuel, JP-4, JP-8, High Density Fuel, Great Plains Gasification Plant, Crude Phenols, Naphtha, Tar Oil, Linear Programming, Economics. (mgn)		
FIELD	GROUP	SUB-GROUP			
21	21	07			
05	05	03			
19. ABSTRACT (Continue on reverse if necessary and identify by block number) Amoco and Lummus Crest have developed seven cases for upgrading by-product liquids from the Great Plains Coal Gasification plant to jet fuels, and in several of the cases, saleable chemicals in addition to jet fuels. The analysis shows that the various grades of jet fuel can be produced from the Great Plains tar oil, but not economically. However, the phenolic and naphtha streams do have the potential to significantly increase (on the order of \$10-15 million/year) the net revenues at Great Plains by producing chemicals, especially cresylic acid, cresol, and xlenol. The amount of these chemicals, which can be marketed, is a concern, but profits can be generated even when oxygenated chemical sales are limited to 10 percent of the U.S. market. Another concern is that while commercial processes exist to extract phenolic mixtures, these processes have not been demonstrated with the Great Plains phenolic stream.					
20. DISTRIBUTION/AVAILABILITY OF ABSTRACT ☒ UNCLASSIFIED/UNLIMITED ☐ SAME AS RPT. ☐ DTIC USERS			21. ABSTRACT SECURITY CLASSIFICATION Unclassified		
22a. NAME OF RESPONSIBLE INDIVIDUAL William E. Harrison III			22b. TELEPHONE (Include Area Code) 513-255-6601		22c. OFFICE SYMBOL AFWAL/POSF

SUMMARY

Amoco and Lummus Crest are contracted with DOE to develop an upgrading scheme for the liquid by-products (tar oil, phenols, and naphtha) produced by the Great Plains gasifiers. These streams are currently burned in the utility boilers and steam superheaters in the Great Plains plant. Task 1 of the contract is complete and the results are reported here. The objective of this task is to develop preliminary economics for seven different upgrading schemes with product slates ranging from maximum jet fuel (JP-4, JP-8, and JP-8X) production to chemical production (cresylic acid, phenol, cresol, xylenol, benzene, toluene, and xylene). A linear program has been developed to evaluate the various upgrading alternatives.

The analysis shows that the various grades of jet fuel can be produced from the Great Plains tar oil, but not economically. However, the phenolic and naphtha streams do have the potential to significantly increase (on the order of \$10-15 million/year) the net revenues at Great Plains by producing chemicals, especially cresylic acid, cresol, and xylenol. The amount of these chemicals, which can be marketed, is a concern, but profits can be generated even when oxygenated chemical sales are limited to ten percent of the U. S. market. Another concern is that while commercial processes exist to extract phenolic mixtures, these processes have not been demonstrated with the Great Plains phenolic stream.

The revenues from chemical sales are sufficient to subsidize the production of jet fuel from the tar oil stream. The economics for these cases are very sensitive to the cost of the replacement fuel for the utility boilers. Replacement fuel costs cannot exceed approximately \$3.00/million Btu if a ten percent real rate of return is to be realized on the capital invested in the upgrading plant.

It should be stressed that these results are preliminary. Numerous assumptions, which need to be verified in subsequent tasks or additional projects, have been made in arriving at the above conclusions.



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FOREWORD

In September 1986, the Fuels Branch of the Aero Propulsion Laboratory at Wright-Patterson Air Force Base, Ohio, commenced an investigation of the potential for production of jet fuel from the liquid by-product streams produced by the gasification of lignite at the Great Plains Gasification Plant located in Buelah, North Dakota. Funding was provided to the Department of Energy (DOE) Pittsburgh Energy Technology Center (PETC) to administer the experimental portion of this effort. This report details the effort of AMOCO Oil Company, who as a contractor to DOE (DOE Contract Number DE-AC22-87PC90015), conducted a preliminary analysis of upgrading alternatives for the production of turbine fuels from the Great Plains liquid by-product streams. DOE/PETC was funded through Military Interdepartmental Purchase Request (MIPR) FY1455-86-N0657. Mr. William E. Harrison III was the Air Force Program Manager, Mr. Gary Stiegel was the DOE/PETC Program Manager, and Drs. Bruce Fleming and Mark Furlong were the AMOCO Program Managers.

DISCLAIMER

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SECTION I

INTRODUCTION

The Great Plains Coal Gasification Plant in Beulah, North Dakota, produces about 145 MM SCF/D of substitute natural gas (SNG) from lignite. The plant also produces three liquid by-products: about 2,900 B/D of tar oil, 830 B/D of crude phenols, and 650 B/D of naphtha. These liquids are all products from the devolatilization of lignite in the Lurgi gasifiers. Currently, the by-products are burned in the plant's boilers and superheaters to produce steam. The economic viability of the plant might be improved by producing marketable products, rather than steam, from these by-product liquids. To this end, Amoco and Lummus Crest, under a contract with the United States Department of Energy, are investigating the technical and economic feasibility of converting the by-product liquids to jet fuels and other saleable products. Jet fuels are of particular interest because of the close proximity of Great Plains to several U. S. Air Force bases.

SECTION II

PROJECT OVERVIEW

As shown in Figure 1, this project is divided into five major tasks: Process Concept Definition, Bench Scale Testing, Pilot Plant Testing, Preliminary Process Design and Economics, and Production Run Recommendation. The results of the first task are reported here.

The first task, Process Concept Definition, includes three subtasks: Liquid By-product Analysis, Process Modelling and Design, and Economic Modelling. The first of these subtasks (1.1), By-Product Analysis, involves analytical characterizations of samples of each by-product taken at six-week intervals. The results from this program, which provide an indication of the average quality of each stream and the variability of that quality over time, are an important input to the second subtask (1.2), Process Modelling and Design. Other inputs to the second subtask include limited experimental processing data on the Great Plains by-products by the Western Research Institute (WRI),⁽¹⁾ Amoco's petroleum refining process models and linear programming technology, Lummus' process simulation and design programs and a market analysis of by-products from Great Plains developed by Sinor Consultants.⁽²⁾ In addition, throughout Task 1, ANG Coal Gasification Company provided valuable input and advice on all fronts. The major objective of subtask 1.2 is to produce seven conceptual designs and associated capital and operating costs for facilities to refine the Great Plains by-products. These seven designs are listed in Figure 2. They include designs for maximizing production of each grade of jet fuel (JP-4, JP-8, JP-8X), designs for profitable schemes which produce the various jet fuel grades, and a scheme for maximizing profits. In subtask 1.3 the results generated by Amoco and Lummus are subjected to economic analysis.

The two products from Task 1 are the design and economic results for each of the seven designs and a plan for bench scale testing (Task 2) to confirm any assumptions made in Task 1. Based on the design and economic results from Task 1 and the experimental results from Task 2, the Department of Energy and the Department of Defense will decide on a preferred processing scheme for the

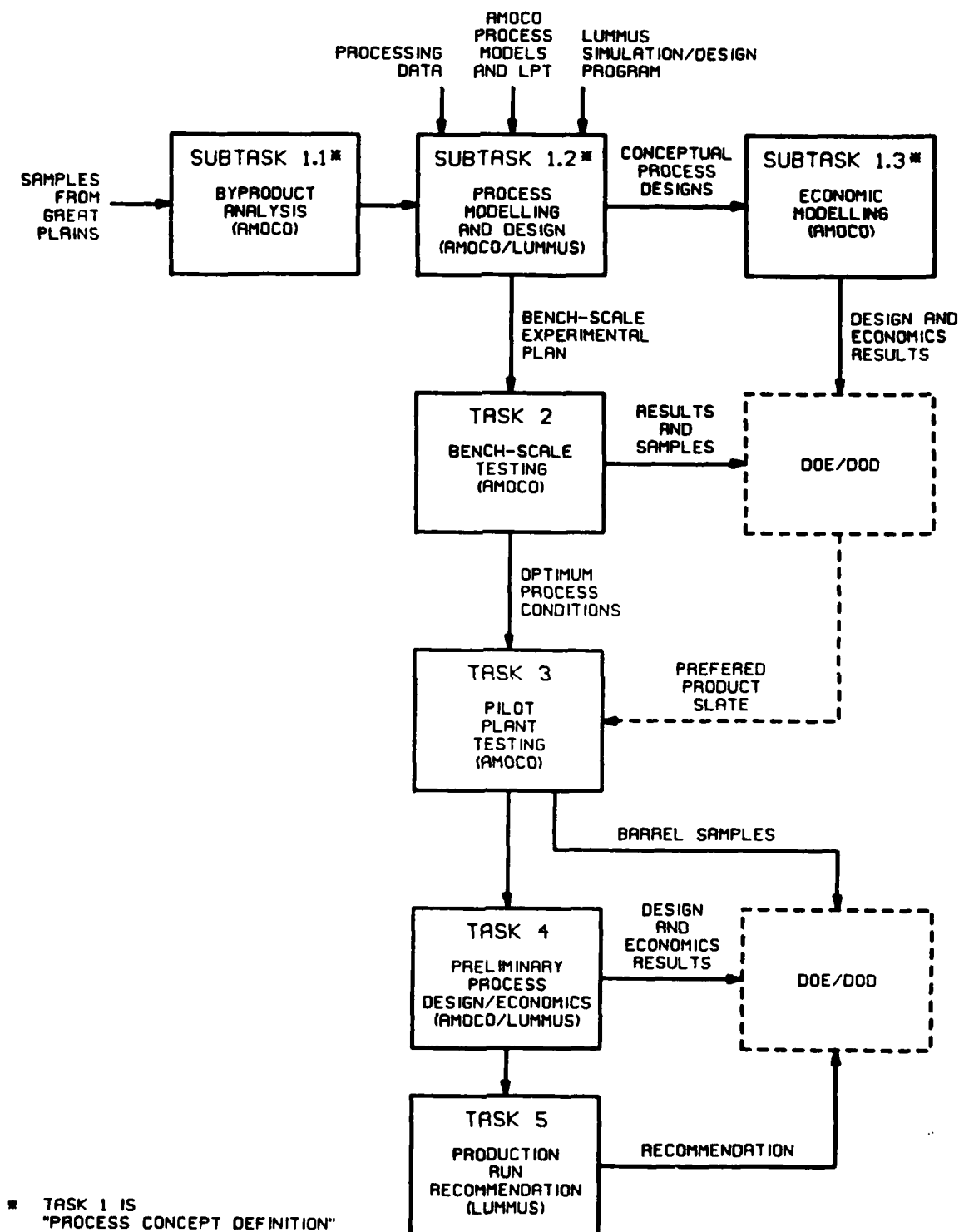


FIGURE 1
PRODUCTION OF JET FUEL
FROM COAL DERIVED LIQUIDS:
AMOCO/LUMMUS ACTIVITIES

THE FOLLOWING DESIGN CASES WILL RESULT FROM ACTIVITIES IN THIS
SUBTASK:

<u>CASE</u>	<u>DESCRIPTION</u>
1	MAXIMUM JP-4 PRODUCTION.
2	PROFITABLE PRODUCT SLATE INCLUDING JP-4.
3	MAXIMUM JP-8 PRODUCTION.
4	PROFITABLE PRODUCT SLATE INCLUDING JP-8.
5	MAXIMUM JP-8X PRODUCTION.
6	PROFITABLE PRODUCT SLATE INCLUDING JP-8X.
7	MAXIMUM PROFIT.

NOTE: CASES 2, 4, 6, AND 7 REQUIRE LINEAR PROGRAMMING TECHNOLOGY.

FIGURE 2

SUBTASK 1.2: PROCESS MODELING AND DESIGN CASE SUMMARY

Great Plains liquids. Amoco will then carry out pilot plant testing (Task 3) to confirm the process design and to provide barrel quantities of product for testing by the government. The pilot plant results will be used by Amoco and Lummus to develop a preliminary process design (Task 4) for a plant to upgrade the liquid by-products at Great Plains. Finally, in Task 5, Lummus will locate existing facilities where the processing scheme can be carried out on a scale sufficient to provide jet fuel for aircraft testing.

SECTION III

SOURCE OF BY-PRODUCT LIQUIDS

Tar oil, crude phenols, and naphtha are produced at the Great Plains Gasification Plant; a schematic of the plant is shown in Figure 3. The plant currently produces about 145 MMSCFD of synthetic natural gas (SNG) from North Dakota lignite. The SNG is composed almost entirely of methane, which is derived mostly from synthesis gas ($H_2 + CO$) produced in the Lurgi Mark IV gasifiers and methanated in downstream reactors. Some methane is produced by devolatilization of the coal in the gasifiers. The liquid by-products (tar oil, phenolics, and naphtha) are produced during lignite devolatilization in the gasifiers.

The tar oil and phenolics are condensed from the product gas along with water vapor to form a gas liquor. This condensation takes place in heat exchangers located in the gasifier quench, shift converter, gas cooling, and Rectisol units. The liquor is routed to the gas liquor separation unit, where the tar oil is recovered by gravity separation. The heaviest portion of the tar oil, which contains about 20 percent coal dust, is recycled to the gasifiers. The recycle rate of this "dusty tar" is about 1800 B/D. The remaining tar oil, which contains 2-6 percent dust, is produced at a rate of 2900 B/D. The phenolics are recovered from the gas liquor by extraction with isopropyl ether in the Phenolsolvan unit. The resulting crude phenol stream, which is produced at a rate of about 830 B/D is composed mostly of phenol, cresol, and xlenol, with the remainder being water and neutral oils. The naphtha is condensed from the gasifier raw gas by contacting the stream with cold methanol in the Rectisol unit. The naphtha is produced at a rate of 650 B/D.

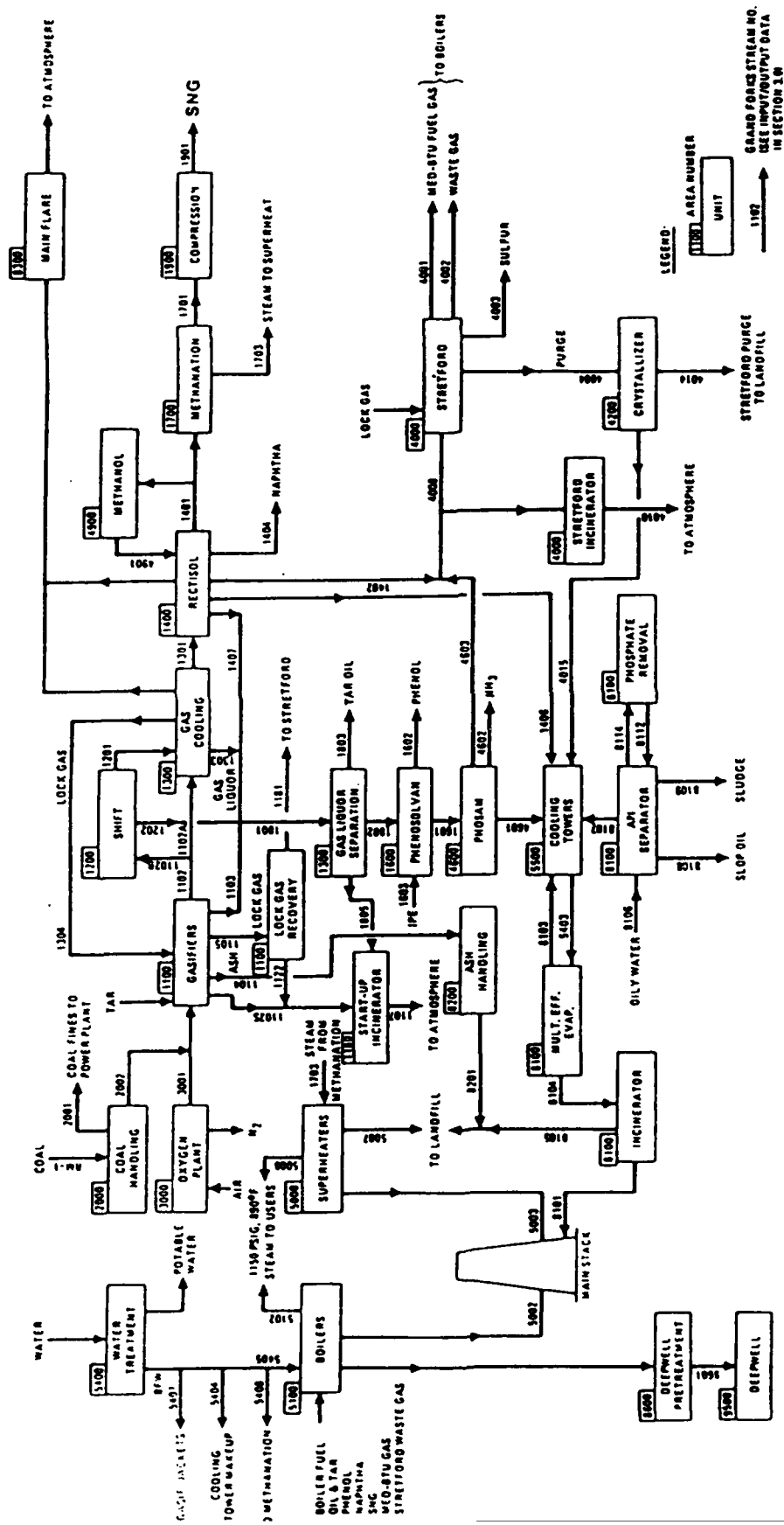


FIGURE 3. GREAT PLAINS GASIFICATION PROCESS BLOCK FLOW DIAGRAM

SECTION IV

TASK 1 RESULTS

1. Subtask 1.1 By-Product Analysis

Each of the three by-product streams from the Great Plains Coal Gasification Plant have been sampled at six to eight week intervals since May of 1987 to monitor seasonal and operational variations in quality and to characterize them as feedstocks for the processes under consideration. This program will continue until mid-1988. These data, with hydrotreating kinetics results from Task 2, will be utilized in Task 4 to prepare a preliminary design of the proposed commercial upgrading/refining facility, integrated into the existing Great Plains plant.

a. Tar Oil

The tar oil stream is the most viable of the by-product streams as a feedstock for the production of jet fuel. ASTM distillations and analyses characterizing this stream are shown in Figure 4 and Table 1. All distillation results have been obtained according to ASTM Method D-2887 with an aromatic standard, except for one D-86, and one D-1160 corrected to D-86. D-86 and D-1160 results are reported on a volume basis and D-2887 on a weight basis. While volume results cannot be rigorously converted to weight without a density profile of the material, the distillations show fairly constant composition among the samples tested. About 2-6 percent of the material boils at less than 300°F and 8-20 percent boils above 800°F on a weight basis.

Previous analyses at the University of North Dakota Energy Research Center (UNDERC)⁽³⁾ by Wilson and co-workers showed the tar oil contained 90 to 95 percent aromatics, the remainder being paraffins. WRI's⁽¹⁾ hydrogenation data show that significant amounts of jet fuel can be made from this stream, but only at severe processing conditions (2,000 psig) and high hydrogen

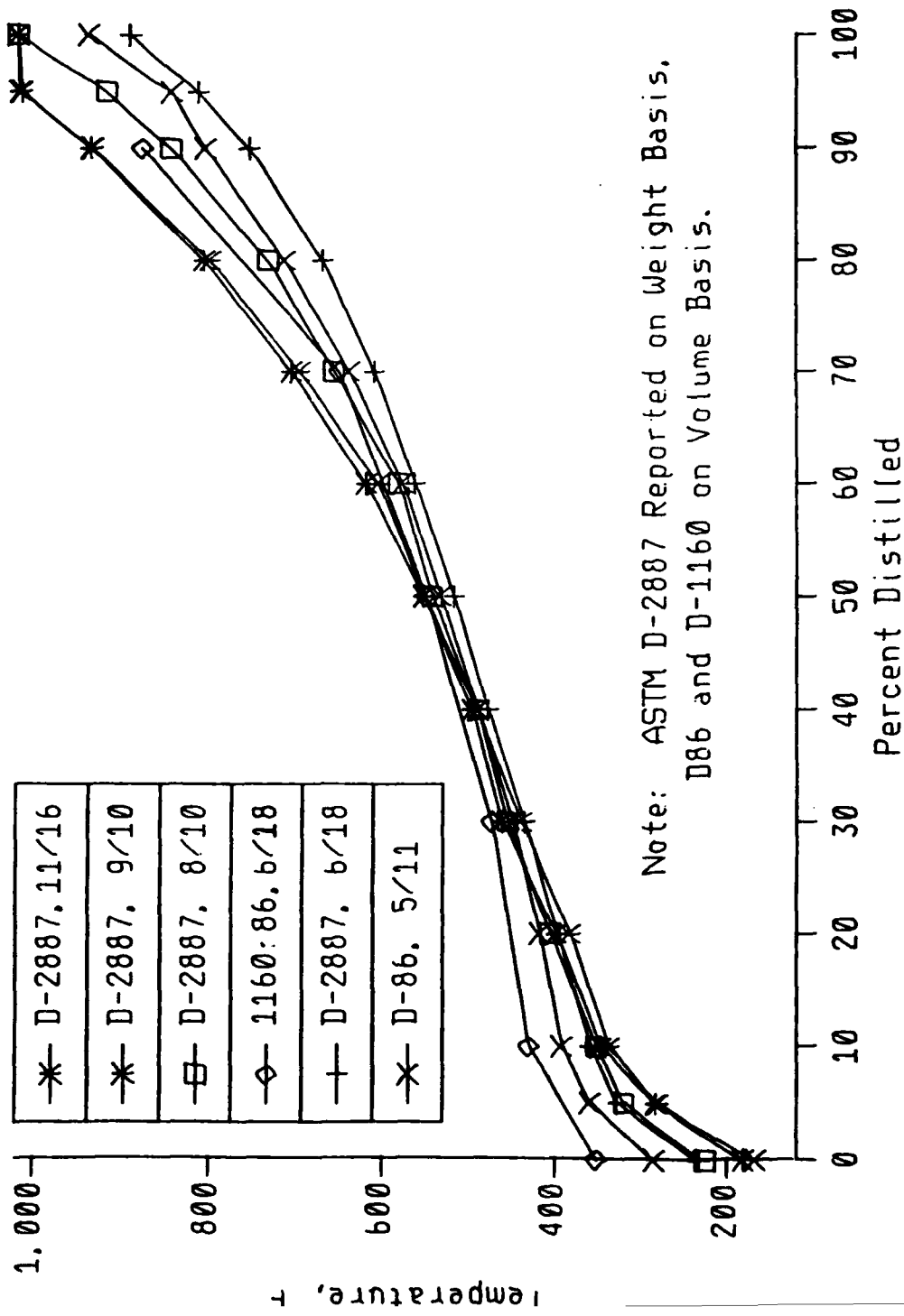


FIGURE 4
GREAT PLAINS COAL LIQUIDS UPGRADING
TAR OIL DISTILLATION RESULTS

TABLE 1

GREAT PLAINS COAL LIQUIDS UPGRADING

FEEDSTOCK CHARACTERIZATION -- TAR OIL

Sample Description:	5/11/87	6/18/87	8/10/87	9/25/87	11/16/87
Water, wt. %	2.04	11.38*	2.30	2.69	1.41
Chlorides, ppm (total/organic)	8/4	--	4/4	2.2	1.0
Elemental Analysis, Dry Basis:					
Carbon, wt. %	83.83	84.80	84.50	84.76	84.19
Hydrogen, wt. %	8.7	9.9	8.9	8.8	8.7
Nitrogen, ppm	8,932	5,495	6,612	6,813	8,297
Sulfur, ppm	4,083	4,006	4,094	4,316	4,730
Oxygen, ppm (diff.)	61,198	43,986	55,210	52,996	57,328
Molar H/C Ratio	1.24	1.38	1.26	1.24	1.23
¹³ C NMR, wt. %CA	61.4	--	--	--	63.8
API Gravity	6.6	8.0	7.4	7.5	7.8
Specific Gravity	1.025	1.014	1.019	1.018	1.016
Viscosity, cp @ 25C	185.1	--	--	--	--
Cloud Point, F	70	--	--	--	--
Free Oxide, wt. %	0.0	0.1	0.0	0.1	0.0
Filtered Solids, wt. %	0.25	0.60	--	0.25	.16/.20**
PSD (Microtrac), %					
<4 microns	17.6	18.3	--	12.1	7.4
4.4-6.6	17.6	12.7	--	10.3	6.8
6.6-9.4	21.2	15.6	--	13.9	9.0
9.4-13	16.8	10.3	--	16.8	10.4
13-19	17.4	12.0	--	20.3	13.4
19-27	8.4	19.1	--	17.4	11.0
>27 microns	0.8	12.0	--	8.7	41.5

Contaminated Sample
Fresh/Aged 4-Wk. @ 150F

consumption (3000-4000 SCFB). Most of the hydrogen consumed goes to saturate aromatics to meet jet fuel specifications (less than 25 percent aromatics).

Water may damage hydrotreating catalysts, and the content of hydrotreated products of materials boiling below about 300°F which may be present in jet fuels is limited. Therefore, water, along with pyridines and other heteroatomic materials boiling at less than 300°F, should be removed by distillation before hydrotreating to decrease hydrogen requirements.

Bench-scale distillations of tar oil with and without addition of toluene as an entrainer demonstrated that most of the water can be removed without use of an entrainer, as shown in Table 2.

Storage of the tar oil sample of 11/16/87 for four weeks at 150°F resulted in no significant changes in ASTM distillation and solids content, shown in Figure 5 and Table 1, respectively.

b. Phenols

The phenolic stream is characterized in Table 3. This stream is composed almost entirely of single-ring hydroxyaromatics, as shown by its high oxygen content of 16-20 weight percent. According to WRI,⁽¹⁾ the crude phenol can be easily hydrogenated to produce highly naphthenic JP-4 blendstock, but the hydrogen consumption is very high, around 5,000 SCFB.

Distillation results for the phenol stream, shown in Figure 6, indicate very good agreement among the samples tested. ASTM Method D-86 is used for all phenol samples. The analytical results, listed in Table 3, also show little variability with a hydrogen-to-carbon ratio of 1.21 to 1.24.

As with the tar oil, the 300°F- fraction of the phenolic stream (about 10 percent of the total) should not be hydrotreated because the water may damage catalysts, and the content of low boiling compounds is limited for jet fuel production. The phenolics stream is also stable, as shown by Figure 5.

TABLE 2
GREAT PLAINS COAL LIQUIDS UPGRADING
TAR OIL BENCH-SCALE DISTILLATIONS RESULTS

METHOD	WATER, WL. %
KARL FISHER ANALYSIS, AS-RECEIVED TAR OIL	1.41
AZEOTROPIC DISTILLATION, TOLUENE	1.28
VACUUM DISTILLATION, NO ENTRAINER	1.21

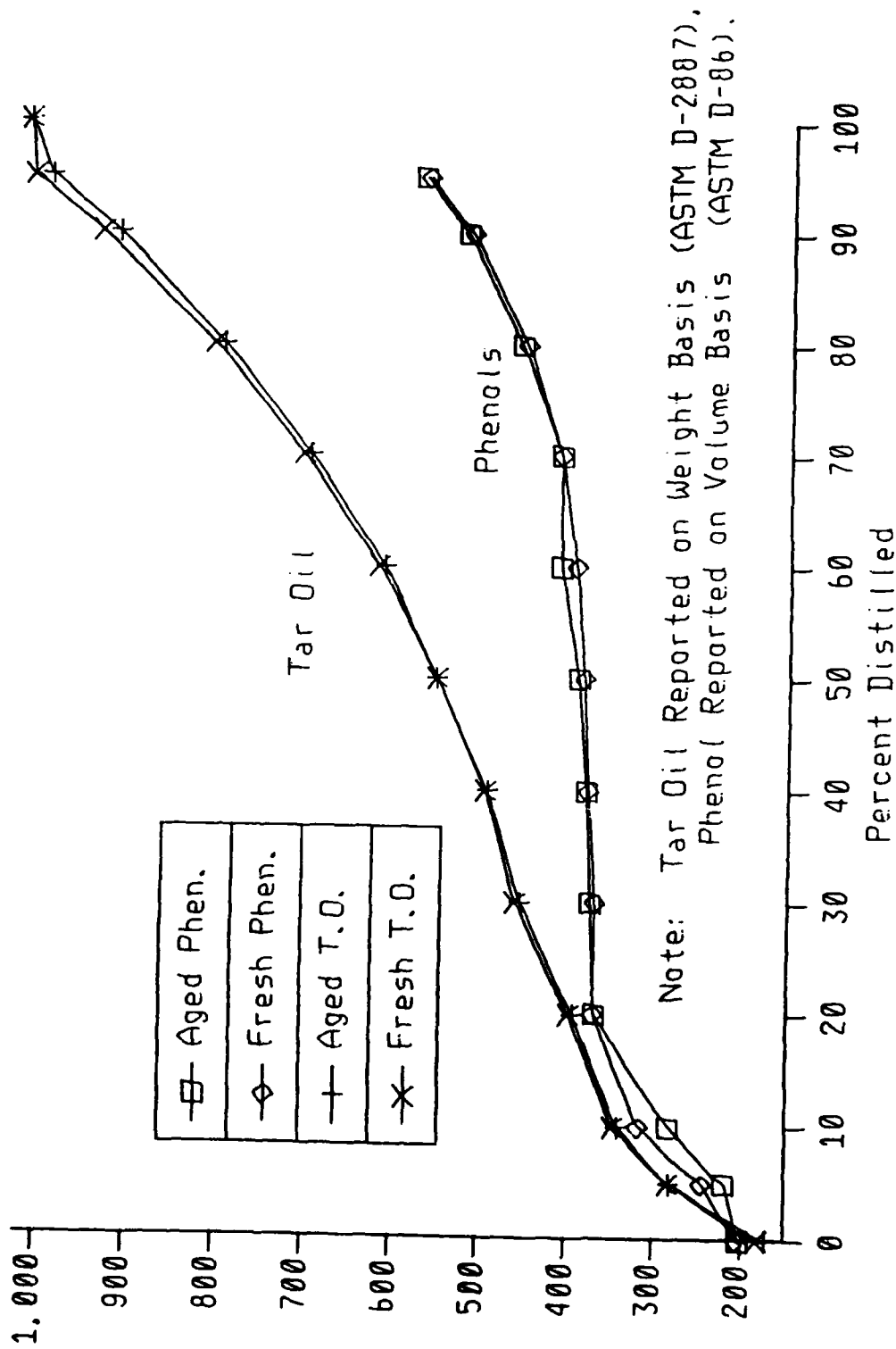


FIGURE 5
GREAT PLAINS COAL LIQUIDS UPGRADING
SIMULATED DISTILLATION RESULTS
BEFORE/AFTER 4-WEEKS STORAGE @ 150 F

TABLE 3

GREAT PLAINS COAL LIQUIDS UPGRADING
FEEDSTOCK CHARACTERIZATION -- PHENOLS

Sample Description:	5/11/87	6/18/87	8/10/87	9/25/87	11/16/87
Water, wt. %	--	--	5.46	5.86	5.56
Phenolides, ppm total/organic)	--	4/5	5/3	6.8	<1
Elemental Analysis, Dry Basis:					
Carbon, wt. %	72.88	72.38	76.76	76.76	75.89
Hydrogen, wt. %	7.6	7.4	7.8	7.8	7.7
Nitrogen, ppm	5,360	5,060	4,876	4,748	4,892
Sulfur, ppm	780	570	751	914	837
Oxygen, ppm (diff.)	189,060	197,070	148,593	148,789	158,503
Atomic H/C Ratio	1.24	1.21	1.21	1.21	1.21
C NMR, wt. %CA	82.6	--	--	--	--
Density	1.6	2.3	1.4	0.9	0.1
Specific Gravity	1.063	1.058	1.065	1.069	1.075
Viscosity, cp @ 25C	15.3	--	--	--	--
Cloud Point, F	-40	--	--	--	--

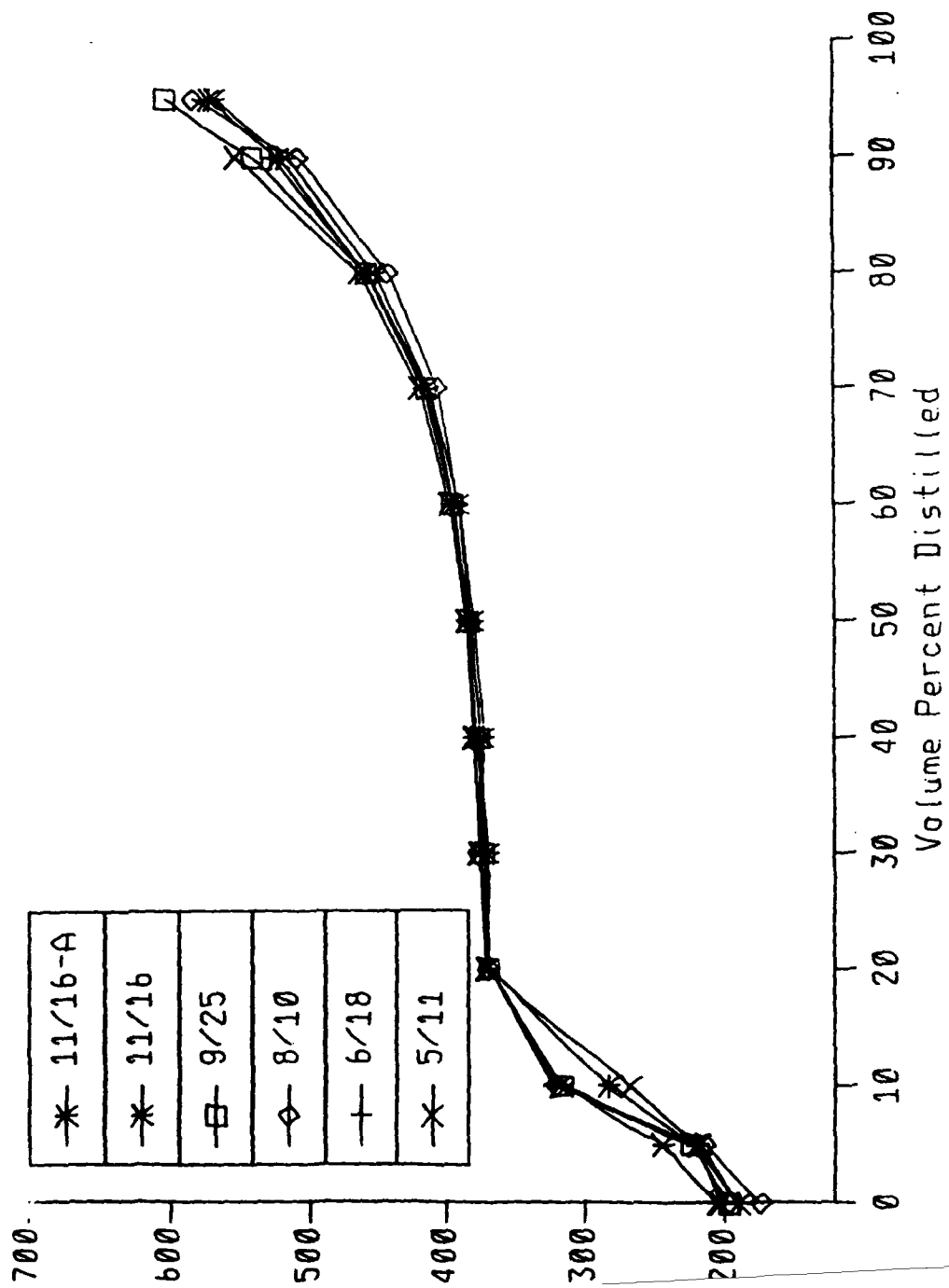


FIGURE 6
GREAT PLAINS COAL LIQUIDS UPGRADING
PHENOLS DISTILLATION RESULTS
(Method: ASTM D-86)

c. Naphtha

The naphtha stream is characterized by analyses shown in Table 4 and distillation results in Figure 7. Comparison of as-received and caustic-extracted samples shows significant decreases in carbon-to-hydrogen ratio, sulfur, and nitrogen for extracted naphtha. As-received samples contain significant amounts of methanol, acetone, and methylethylketone from the Great Plains Rectisol unit, and the presence of thiols, mercaptans, and thiophenes. These produce extremely noxious odors when opened. Resulting safety and environmental concerns have limited the number of naphtha analyses and distillations performed. The boiling range of the naphtha stream is too low to produce significant amounts of jet fuel. Aromatics can be recovered from this stream, but only after hydrotreatment to reduce sulfur and nitrogen levels.

2. Subtask 1.2 Process Modelling and Design

The seven design cases for which conceptual process designs have been developed are listed in Figure 2. The first six cases, which are required by the contract, involve making jet fuel either at maximum production rate or with co-products such that the total product slate is profitable. The seventh case has been added by Amoco as a natural extension of the project. This case drops the requirement of production of jet fuel and produces the most profitable product slate.

TABLE 4

GREAT PLAINS COAL LIQUIDS UPGRADING

FEEDSTOCK CHARACTERIZATION -- NAPHTHA

Sample Description:	5/11/87	5/11/87 Extract	6/18/87	6/18/87 Extract	8/10/87
Water, wt%	0.56	--	0.42	0.11	0.51
Elemental Analysis, Dry Basis:					
Carbon, wt%	84.22	86.33	84.46	86.30	84.85
Hydrogen, wt%	10.00	9.57	9.97	9.93	9.94
Nitrogen, ppm	2,062	630	2,089	1,321	1,970
Sulfur, ppm	17,397	12,600	16,871	861	18,394
Oxygen, ppm (diff)	38,365	27,770	36,674	35,459	31,702
Atomic H/C Ratio	1.41	1.32	1.41	1.37	1.40
¹³ C NMR, wt% CA	65.9	--	--	--	--
API Gravity	38.9	39.5 40.0	39.8	53.8	
Specific Gravity	0.830	0.827	0.825	0.826	0.764
Viscosity, cP @ 25C	0.5	--	--	--	--
PONA Analysis, Liq. Vol%					
Paraffins	23.1	7.4	8.3	--	--
Naphthenes	12.2	3.9	4.5	--	--
Olefins, noncyc/cy	--	14.4/9.7	15.6/7.7	--	--
Aromatics	64.6	64.6	63.9	--	--

TABLE 4 (concluded)
 GREAT PLAINS COAL LIQUIDS UPGRADING
 FEEDSTOCK CHARACTERIZATION -- NAPHTHA

Sample Description:	5/11/87	5/11/87 Extract	6/18/87	6/18/87 Extract	8/10/87
Light Ends, wt%	(bomb)				
C1	--	--	--	--	0.0
C2	--	--	--	--	0.0
nC3	0.0	0.1	0.0	--	0.0
C3=	--	--	--	--	--
nC4	0.1	0.2	0.1	--	0.1
iC4	0.0	0.0	0.0	--	0.0
C4=	0.438	1.064	--	0.725	-- 0.886
nC5	0.469	0.603	--	0.217	-- 0.259
iC5	0.158	0.196	--	0.175	--
C5 cyc.	0.241	0.280	0.253	--	0.291
C5=	2.258	2.735	2.046	--	3.003
C5==	3.507	4.099	2.887	--	2.932
Misc. Components, wt%					
Methanol	2.27	--	2.47	--	1.82
Acetone	6.32	--	--	--	--
MEK	3.32	--	--	--	--
Benzene	45.62	--	46.04	--	46.30
Toluene	17.93	--	15.99	--	16.54
Xylenes (incl. EB)	3.25	--	3.60	--	3.86
C5==	3.507	4.099	2.887	--	2.932
Misc. Components, wt%					
Methanol	2.27	--	2.47	--	1.82
Acetone	6.32	--	--	--	--
MEK	3.32	--	--	--	--
Benzene	45.62	--	46.04	--	46.30
Toluene	17.93	--	15.99	--	16.54
Xylenes (incl. EB)	3.25	--	3.60	--	3.86

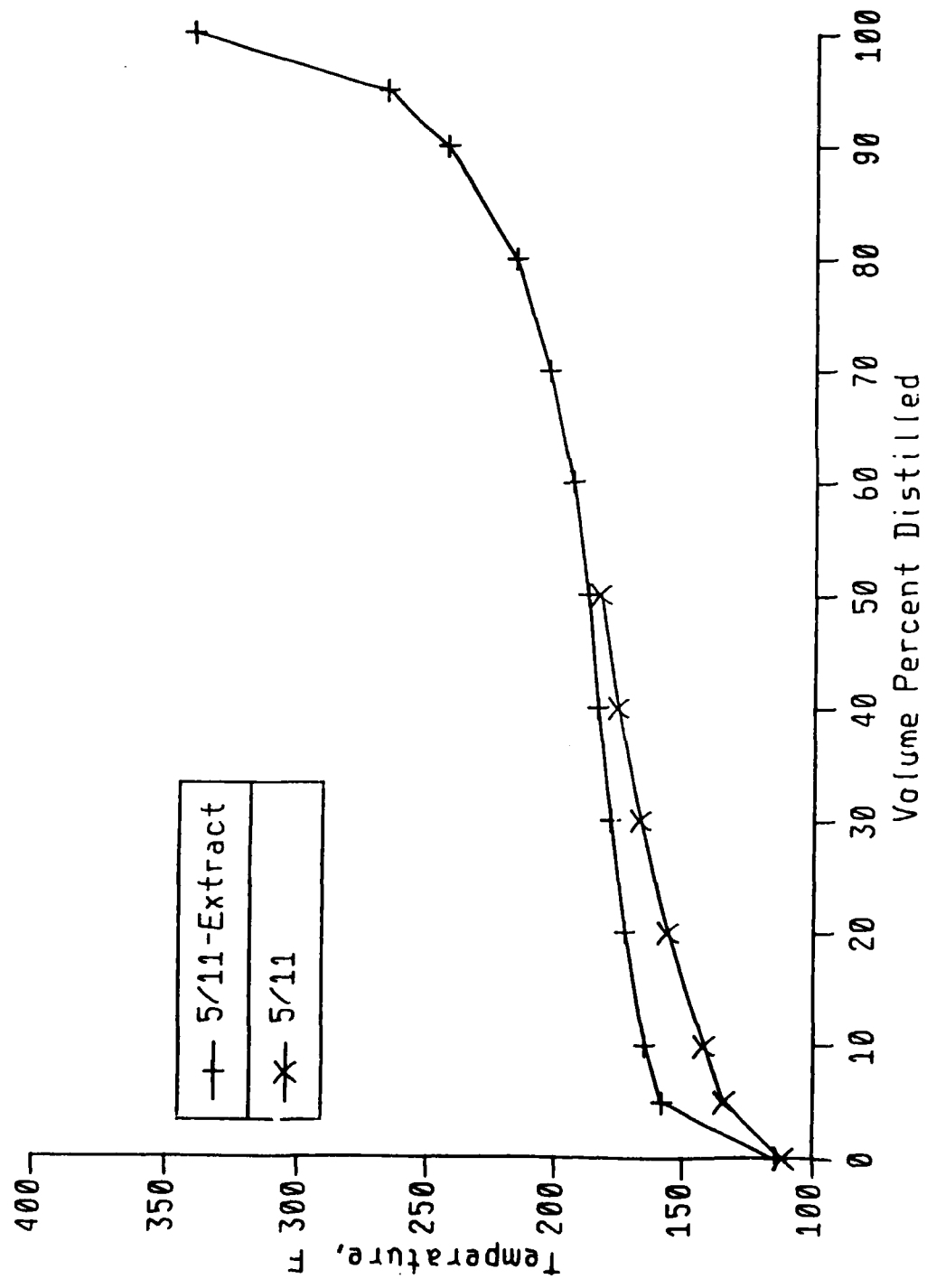


FIGURE 7
GREAT PLAINS COAL LIQUIDS UPGRADING
NAPHTHA DISTILLATION RESULTS
(Method: ASTM D-86)

Many alternative routes to upgrade the by-product streams can be proposed. While it may be easy to pick the route that maximizes jet fuel production, the routes that produce a profitable product slate with jet fuel, or the maximum profit product slate are difficult to determine without using a computerized approach which can consider many alternatives for the profitable cases.

a. Technical Approach

To evaluate all possible cases, Amoco has developed a linear program for the Great Plains by-products. This LP has numerous inputs, including:

1. Amoco's proprietary linear programming technology (including refinery process models).
2. The results of the Western Research Institute scoping hydrotreating study. ⁽¹⁾
3. The price structure for various oxygenated chemicals from the Sinor report. ⁽²⁾
4. Amoco's price structure for hydrocarbon products and feeds in the North Dakota area.
5. Information on processes to upgrade oxygenated chemicals. ^(4,5)

The Great Plains LP contains 1,450 equations and approximately 2,400 variables. The equations describe process and utility constraints, mass and energy balance constraints, blending options, capital and operating costs.

Linear programming is used by Amoco to help guide research and refining planning. It is a methodology to rigorously analyze the effects of process

technology, feeds, and products on refinery configurations. Amoco's linear program technology is based on Amoco's extensive commercial petroleum refining experience. It can be used to evaluate additions and revamps of existing refineries or design a new refinery (utilities available), as would be the case for Great Plains.

Amoco's linear programming technology has an extensive data base. It includes process models; raw material and product specifications, and prices; capital investment, operating cost and utility requirement correlations; and product blending algorithms. The process models are based on Amoco's technology wherever possible, but all technologies are commercially proven and licensable. The data base is used in conjunction with a refinery and petrochemical modeling system that draws on the data base to construct linear program models, which are solved consistent with whatever constraints the user chooses to apply. In essence, a tailor-made LP model is developed for each new problem. The user may specify virtually any combination of feed material and/or product volumes, specifications, and prices. Alternatively, or in addition, refinery configurations can be specified completely, in part, or not at all. Optimal solutions are developed for each variable not specified. This affords a great deal of flexibility in analyzing a wide variety of refining environments and refinery configurations. The LP analysis also considers and optimizes process interactions and product blending effects.

Table 5 lists the processes which are included in the linear program developed for Great Plains liquid by-product upgrading. The major process blocks are also shown schematically in Figure 8. Several processes use standard petroleum refining technology. Most other processes (denoted by (1) in Table 5) have been adapted from standard petroleum technology by adjusting for the different feedstock inspections of the Great Plains streams. Where interpolation between literature or proprietary data points is not possible, proprietary Amoco process simulation models are used to predict the effects of feedstock inspections on process operations. These simulations have been developed from data using petroleum based feeds, including high-aromatics stocks similar to the Great Plains streams. A few processes (denoted by (2) in Table 5) are critical enough or different enough from standard petroleum

processing that customized models were developed. The development of these models is outlined below.

TABLE 5

**PROCESS BLOCKS INCLUDED IN LINEAR PROGRAM
SIMULATION OF GREAT PLAINS LIQUIDS UPGRADING**

Aromatics Recovery (1)
Butane and Pentane Isomerization
Catalytic Cracking (1)
Cresylic Acid Fractionation (1)
Delayed Coker (1)
Distillate Hydrotreater (NiMo & NiW) (2)
Dynaphen (2)
Gas Oil Desulfurizer
Gasoline Reformer (1)
Hydrocracking for JP-4, JP-8, and Gasoline (1)
Hydrogenation/Saturation (2)
Naphtha Distillation (1)
Naphtha Hydrotreater (2)
Naphtha Sweetening
Phenoraffin/ANG Extraction (2)
Pressure Swing Absorption (PSA) (1)
Product Blending
Propylene, Butylene, and Amylene Alkylation
Propylene Concentration
Propylene and Butylene Polymerization
Sulfur Recovery
Tar Oil Distillation (1)
Utilities Generation

-
- (1) Units which differ in feedstock properties from standard petroleum refining technology, but can readily be adapted to Great Plains stocks.
- (2) Units which differ enough from standard petroleum refining technology to be customized for application to Great Plains stocks (see text for details).

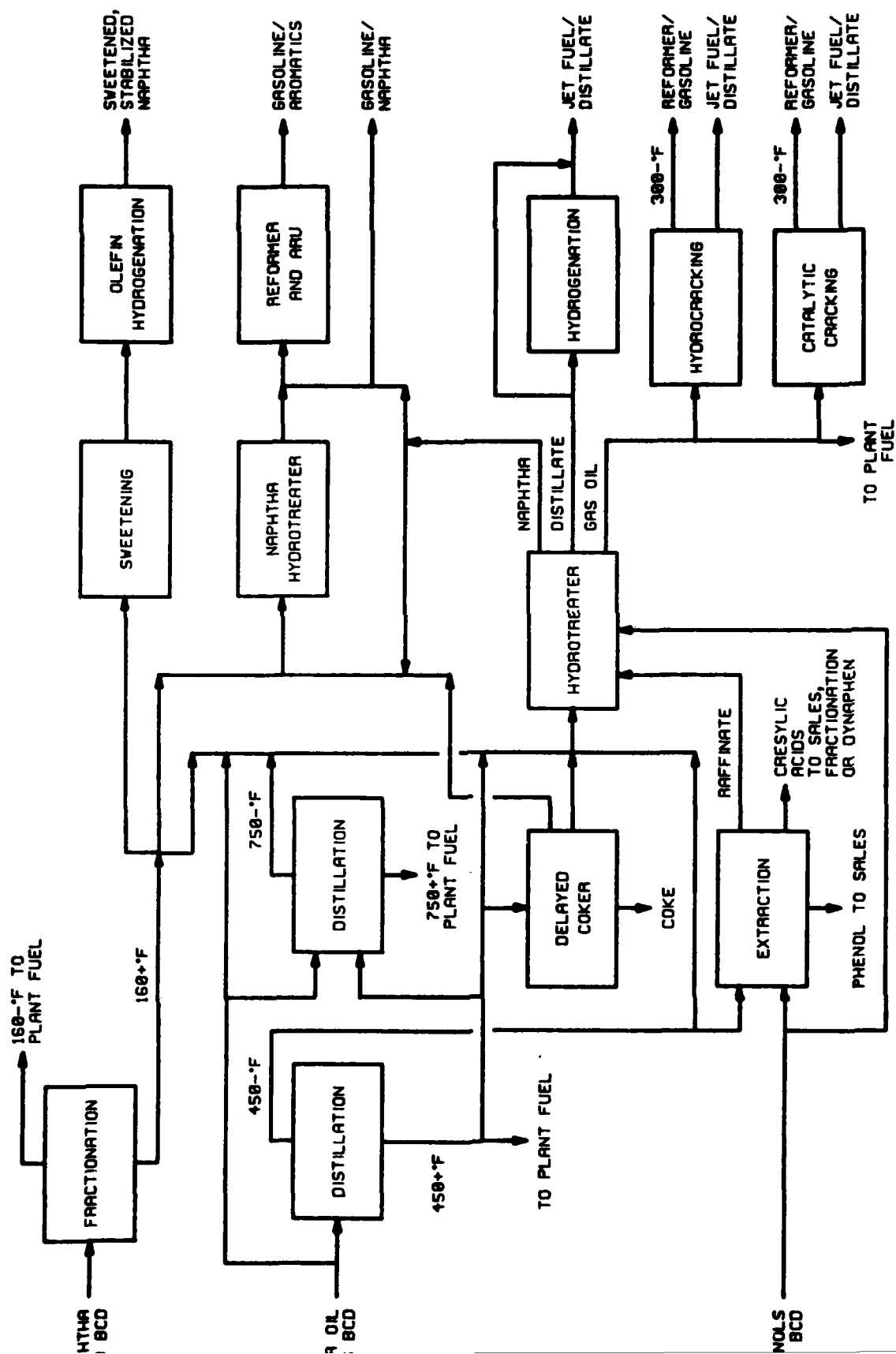


FIGURE 8
LINEAR PROGRAM FLOWSHEET FOR GREAT PLAINS BY-PRODUCT REFINING

The distillate hydrotreating and hydrogenation/saturation processes are modeled using process data from coal-derived feedstocks published by Chevron.⁽⁶⁾ Since aromatics saturation reactions are so important in processing Great Plains streams to meet jet fuels specifications, an aromatics saturation model was developed and reported in Appendix A. The phenol hydrotreating is based on WRI⁽¹⁾ test 87-07-5, adjusted for different feed inspections and adding light ends yields. The hydrotreating results from neutral oils from Phenoraffin extraction are estimated based on a higher aromatics content than the original streams.

An expanded bed technology was selected to hydrotreat the tar oil. This process selection was based on the following advantages for expanded beds, relative to fixed beds, for this feed: (1) Little or no feed preheat is required, (2) Relatively isothermal operation with no large reactor temperature exotherm and relatively simple reactor temperature control requirements, (3) Tolerance to solids in the feed, and (4) Operating flexibility.

The data used for estimating the yields and process conditions for hydrotreating the Great Plains naphtha comes from Chevron's results with SRC-II naphtha.⁽⁶⁾ A proprietary fractionation simulation is used to divide the Great Plains naphtha stream into 160°F- and 160°F+ cuts. The 160°F- cut is used as fuel. The coker naphtha hydrotreating yields are based on proprietary data using delayed coker naphtha from petroleum stocks.

All hydrotreater (naphtha and tar oil) investments and operating costs are based on estimates developed by Lummus, based on their preliminary process designs. Dynaphen processing is based on the literature.⁽⁵⁾ The Dynaphen process, which is still to be confirmed for Great Plains feedstocks by HRI, will partially dealkylate coal liquids to phenol and benzene.

The Phenoraffin process yields are based on published recoveries of pure components adjusted for Great Plains phenolic composition. Phenoraffin is Lurgi's proprietary process for separation and purification of phenols, cresols, and other commodity chemicals from mixed oxygenate streams. A firmer process design and cost basis for Phenoraffin was sought from Lurgi, but they

reportedly needed to do experimental work using Great Plains feedstocks before they would provide this information. As a result, Lummus developed a process design and cost basis for the phenolic extraction based on a dual solvent extraction process under development by ANG.⁽⁴⁾ Henceforth in this report, this process will be referred to as the Phenoraffin/ANG process.

Feed costs and product values are taken from Amoco proprietary North Dakota refinery information and a recent Sinor study.⁽²⁾ Hydrogen cost, based on extraction from Great Plains syngas, is from a recent Burns and Roe study.⁽⁸⁾ Hydrogen is extracted by pressure swing adsorption. The cost of feeds and the value of products used in the linear program are listed in Table 6.

Product specifications for the various grades of gasoline and jet fuel have also been included in the LP data base. These specs are given in Table 7. Specification for the oxygenated chemical products have not been determined for the preliminary evaluation in Task 1. It is assumed that the Phenoraffin/ANG process which produced the various oxygenated chemicals, could meet the required specifications. This assumption should be verified.

b. Conceptual Upgrading Schemes

Using the Great Plains LP, conceptual upgrading schemes have been developed, along with sensitivities to key variables and assumptions. The cases are presented in the order: maximum jet fuels (Case 1:JP-4; Case 3:JP-8; Case 5:JP-8X); maximum profit (Case 7); and profitable jet fuels (Case 2:JP-4; Case 4:JP-8; Case 6:JP-8X).

The maximum jet fuels cases have been estimated by hand-calculation and by the LP. The hand calculation allowed Lummus to begin work before the LP was finished. They also provide a check on the accuracy of the LP. The economics presented below, however, are based on the LP estimated cases. The LP simulations do not exactly match the hand-calculated process schemes because of three reasons. First, the linear program cannot exactly match the non-linear curves which relate yields and qualities to process conditions, because it approximates these curves with straight line segments. Second, the

TABLE 6
COST AND PRICE BASES AND ASSUMPTIONS

<u>Input Streams</u>	<u>Price \$/Bbl</u>	<u>Max. BCD</u>	<u>Fuel Value, \$/MMBtu</u>
Natural gas	13.57	-	2.15
LPG/Propane	7.57	-	2.15
i-Butane	19.11	-	4.98
n-Butane	11.76	-	2.95
GP Naphtha	12.69	660	2.15
GP Phenols	10.43	833	2.15
GP Tar Oil	13.07	2896	2.15
Toluene	38.01	5000	
GP Syngas for H2	1.23/MSCF H2	36810 MSCFD	2.47
<u>Output Streams</u>		<u>Limitation:</u>	
LPG/Propane	7.56	-	
Unleaded Gasoline	23.35	-	
Unleaded Premium	26.29	-	
Sweetened GP Naphtha	25.45	-	
Reformer Feed	24.61	-	
Hydrotreated GP Naphtha	30.00	-	
JP-4	24.19	-	
JP-8	21.84	-	
JP-8X	21.84	-	
Benzene	48.00	-	
Toluene	38.00	-	
Xylene	49.00	-	
Phenol	80.00	-	
o-Cresol	182.00	25	10% U.S. Market
m,p-Cresol	199.00	80	10% U.S. Market
Xylenols	171.00	30	10% U.S. Market
Cresylic Acids	134.00	140	10% U.S. Market
GP Fuel Pool	2.15/MMBTU	25540 MMBTU/CD	
Sulfur	125/LT	-	
Coke	16/T	-	

TABLE 7
PRODUCT SPECIFICATIONS

Jet Fuel Specification	JP-4	JP-8	JP-8X
Boiling Range, °F	160-518	300-572	300-572
Specific Gravity, min.	0.7507	0.775	0.850
" " " , max.	0.8017	0.840	
Sulfur, wt%, max.	0.4	0.3	0.3
Aromatics, min.			10
" " " , max.	25	25	30
Paraffins, max.			10
Naphthenes, min.			70
" " " , max.			90
Reid Vapor Pressure, min.	2.0		
" " " " , max.	3.0		
Flash Point, °F, min.		100	122
Pour Point, °F, max.		-72	-62

Gasoline Specification	Unleaded Regular	Unleaded Premium	Unleaded Midgrade
Road Octane, min.	87	93	89
Reid Vapor Pressure, min.	9.6	9.6	9.6
" " " " , max.	12.7	12.7	12.7
Specific Gravity, max.	0.7669	0.7669	0.7669
Sulfur, wt%, max.	0.1	0.1	0.15

In addition, distillation specifications consistent with ASTM D-439 were employed.

In addition, distillation specifications consistent with ASTM D-1655, MIL-T-5624L, and MIL-T-83133 were employed. Pour point and flash point specifications were met for JP-8 and JP-8X using proprietary non-linear blending techniques.

Also sold but blended from stocks which are produced to required purity specifications are: LPG, sweetened naphtha, reformer feed, hydrotreated naphtha, BTX, phenol, o-cresols, m- and p-cresols, xylenols, cresylic acids, coke, and sulfur. See Table 3.

LP adjusts process conditions in response to optimal costs until process limitations are reached. Thus, the LP design always results in a solution which is limited by some constraint. Finally, a true LP maximum sometimes results in absurd processing choices. For example, the unconstrained LP maximum JP-4 case includes processing the Great Plains naphtha and several other streams, of less than 100 B/D, to get a few more barrels of jet fuel. It does this by building a very small catalytic cracker to get a low gravity jet fuel stock which allows a few more barrels of high gravity stocks to go to jet fuel. The cases reported here externally forbid process options which result in very high costs or very small stream or unit sizes. The prohibited streams are less than 10 percent of the parent stream volumes.

The other four cases have no corresponding hand-calculated solutions to compare with, since they are economic optima. Some external process limits are also imposed on these cases to avoid processing very small streams or building very small units.

Lummus's estimate of investment and operating costs for each case are reported below.

Case 1 - Maximum JP-4.--Figure 9 shows hand-calculated maximum JP-4 case. This case hydrotreats the Great Plains tar oil and phenol streams, while leaving the naphtha for plant fuel needs. The hydrotreated 500°F+ stream is hydrocracked to produce additional jet fuel blending stocks. Combined JP-4 production is 4270 BSD, along with 560 BSD of naphtha stocks. The LP solution, shown in Figure 10, produces about 4 percent more JP-4 than the hand-calculated case. This is within the range expected, due to LP limitations discussed above. Figure 9 reports flows in BSD (barrels per stream day), while Figure 10 reports them in BCD (barrels per calendar day). Using the Great Plains operating factor of 91 percent, BCD values are 9 percent lower than BSD values. Yields of key products, in barrels per calendar day, are compared in Table 8. Both Figures 9 and 10 include the same processes, so the match of these two cases is sufficient for this scoping evaluation.

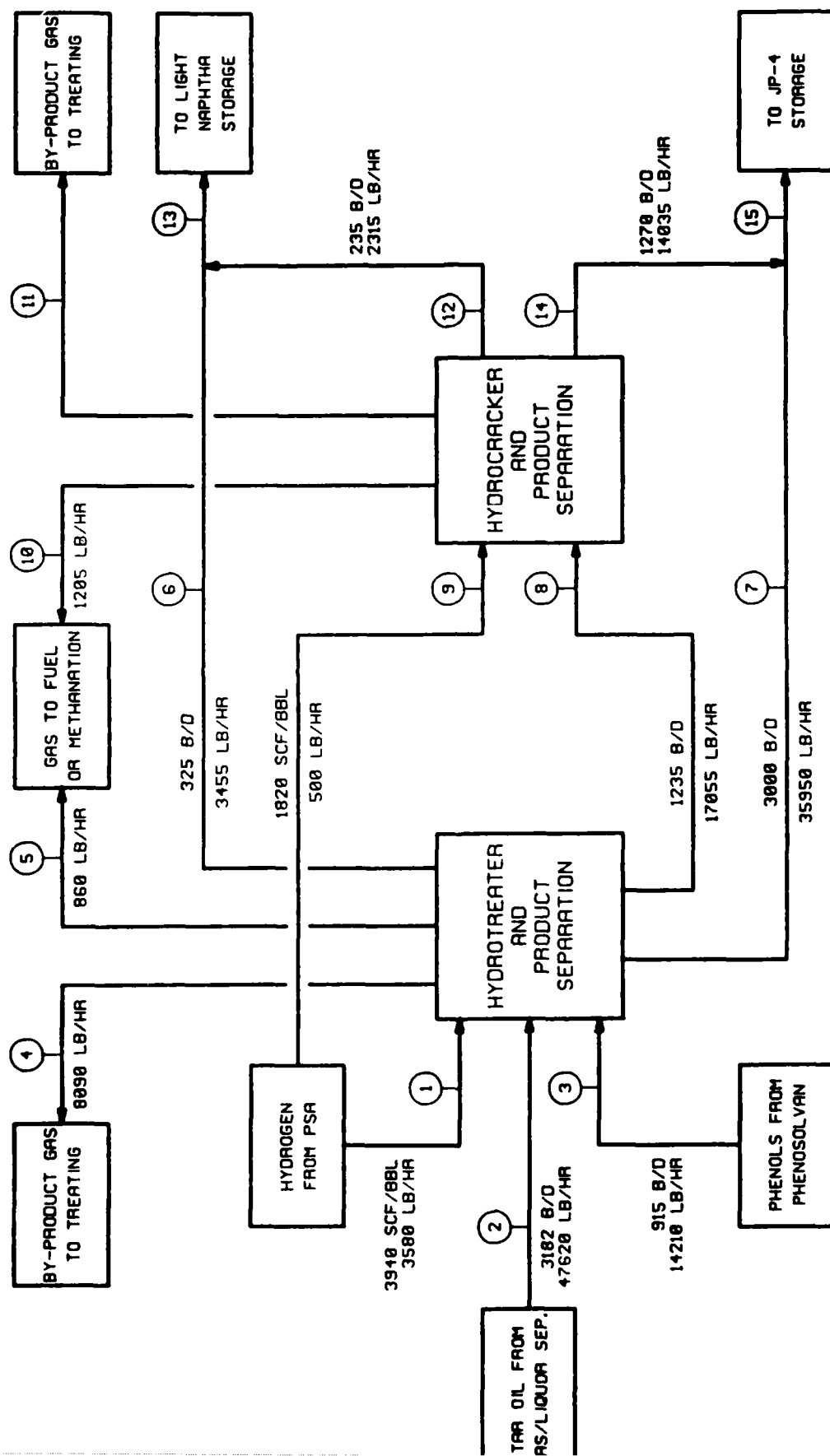


FIGURE 9
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 1: MAXIMUM JP-4 PRODUCTION

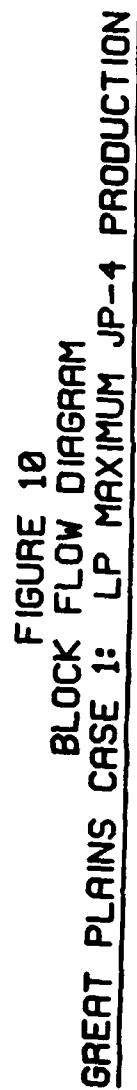


FIGURE 10
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 1: LP MAXIMUM J

TABLE 8

COMPARISON OF MAXIMUM JET FUEL CASES WITH LP SOLUTIONS

<u>Streams</u>	<u>JP-4</u>		<u>JP-8</u>		<u>JP-8X</u>	
	<u>Hand- Calculated</u>	<u>LP</u>	<u>Hand- Calculated</u>	<u>LP</u>	<u>Hand- Calculated</u>	<u>LP</u>
Naphtha, BCD	510	466	1215	1222	505	378
Jet Fuel, BCD	3886	4052	2266	2323	1784	1764
Fuel Oil, BCD	0	0	0	0	804	947
Syngas, MMSCFCD	31.0	30.4	24.0	23.7	13.6	14.3

Case 3 - Maximum JP-8.--Figure 11 shows hand-calculated maximum JP-8 case. This case uses only the tar oil stream to make jet fuel, while leaving both the phenol and naphtha stream for plant fuel needs. The same sequence of hydrotreating and hydrocracking is used as for JP-4 production, but the cut point from the hydrotreater is 550°F instead of 500°F. The linear program maximum JP-8 case, shown in Figure 12, again estimates about 3% more JP-8 than the hand calculated case, as Table 8 shows. Again, the close match of unit operations in Figures 11 and 12 confirms the estimating design basis.

Case 5 - Maximum JP-8X.--Figure 13 shows the hand-calculated maximum JP-8X case. Here too, only the Great Plains tar oil stream is used for jet fuel production. This case first fractionates the tar oil at 750°F and hydrotreats only the 750°F- cut. The hydrotreater products are naphtha, JP-8X, and a 550°F+ stock which is combined with 750°F+ tar oil to provide plant fuel. The high-boiling fractions of the tar oil are too aromatic to make JP-8X. The LP maximum JP-8X case, shown in Figure 14, bypasses some tar oil directly to hydrotreating to save distillation costs. However, the close match of units chosen and overall yields listed in Table 8 again confirms the design basis used. The preferred design basis should be to fractionate all the tar oil, since the incremental cost of this capacity is low and this improves plant operating flexibility.

Case 7 - Maximum Profit.--The maximum profit case is shown in Figure 15. This case processes only the Great Plains naphtha and phenol streams. The tar oil stream is used as plant fuel. The naphtha is fractionated to remove undesirable heteroatoms boiling below 160°F. Improved operation of the existing Great Plains stripper might eliminate this process step, but it is included here if the desired stripper performance cannot be achieved. The 160°F+ naphtha is hydrotreated and then aromatic components (benzene, toluene, and xylene, or BTX) are recovered for sale. The remaining naphtha components are blended to gasoline. A small PSA unit recovers 0.2 MMSCFD of hydrogen from the Great Plains Rectisol gas stream to provide the hydrogen for this hydrotreating step.

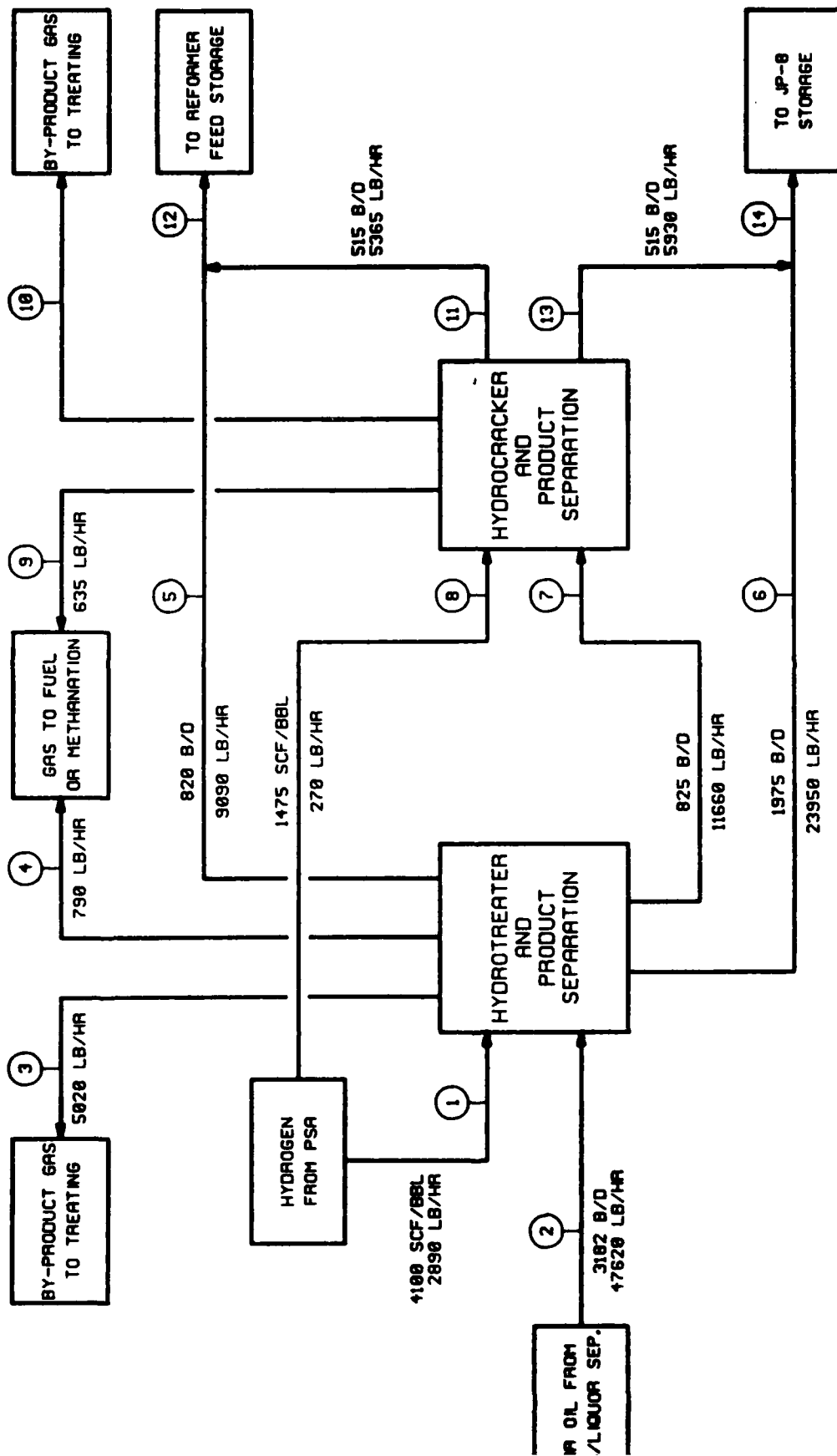


FIGURE 11
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 3: MAXIMUM JP-8 PRODUCTION

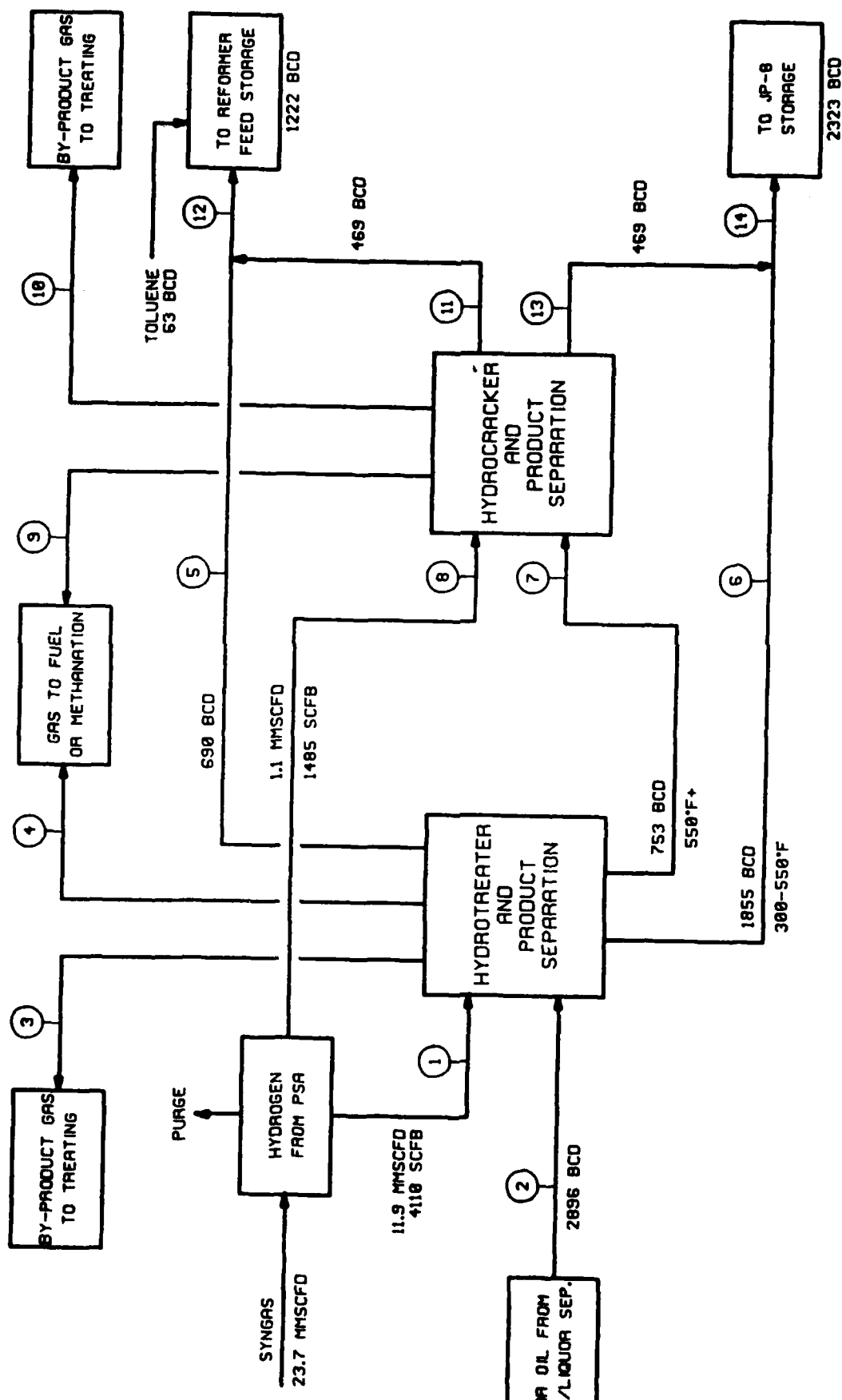


FIGURE 12
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 3: LP MAXIMUM JP-8 PRODUCTION

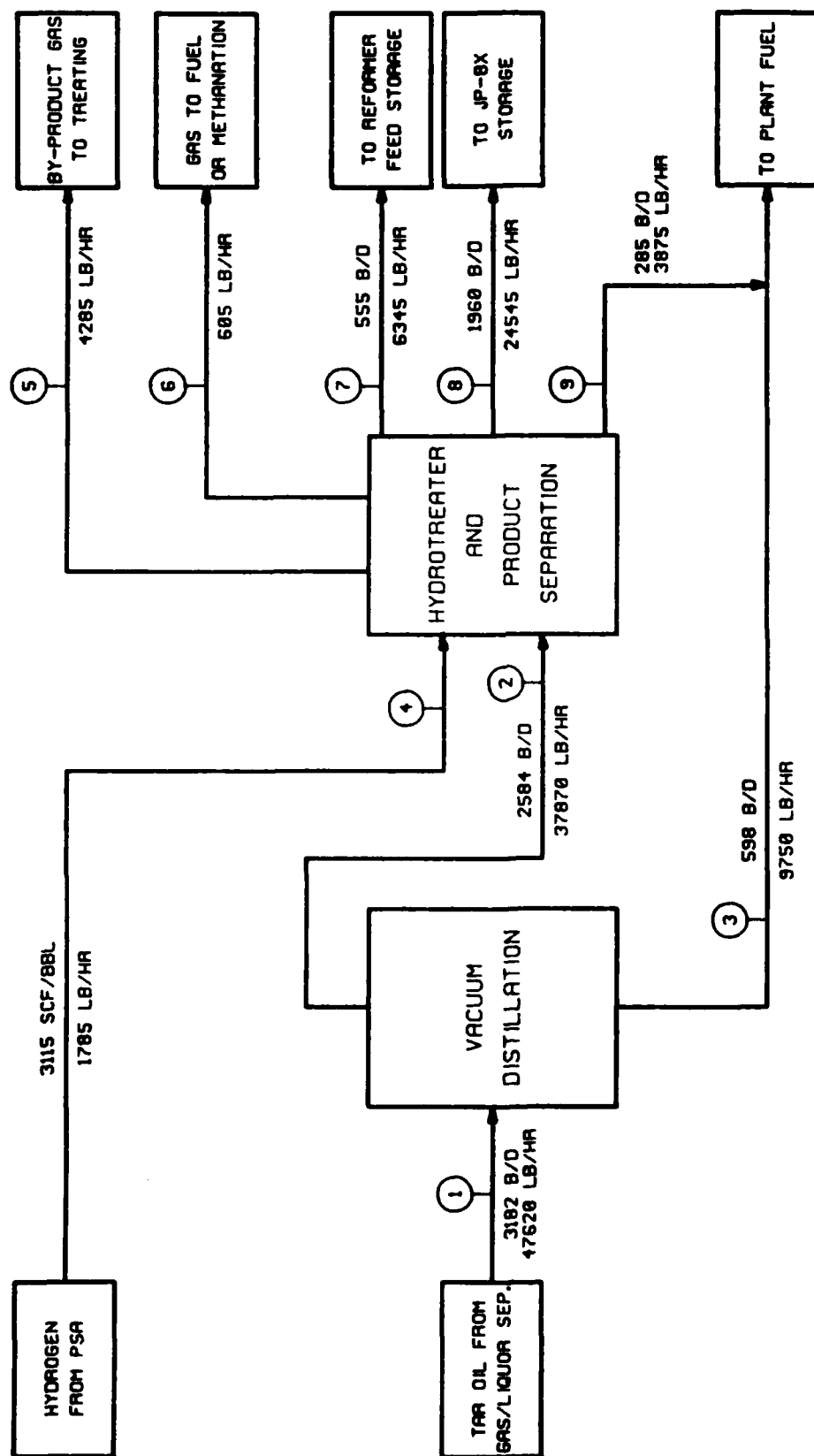


FIGURE 13
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 5: MAXIMUM JP-8X PRODUCTION

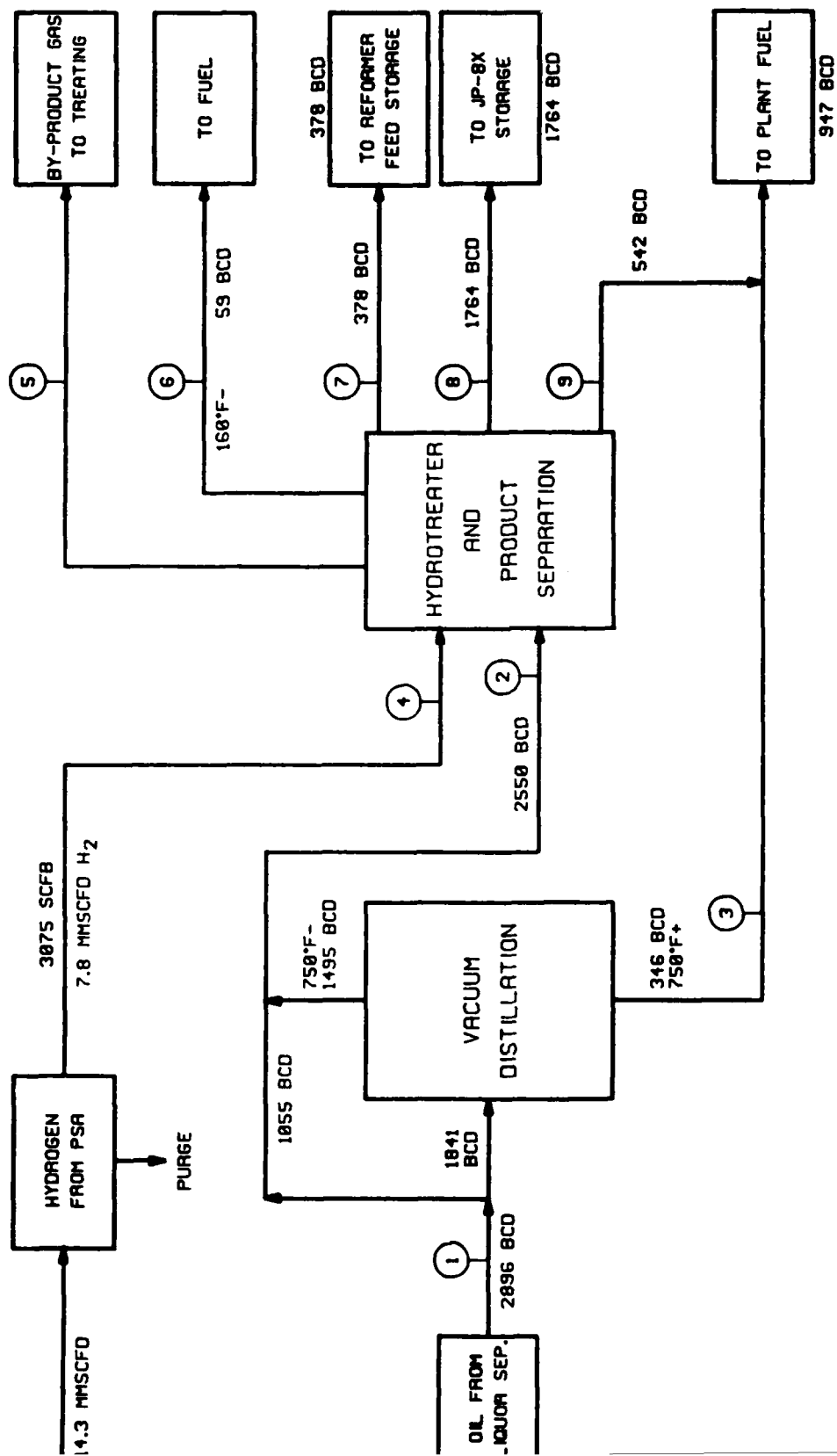


FIGURE 14
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 5: LP MAXIMUM JP-8X PRODUCTION

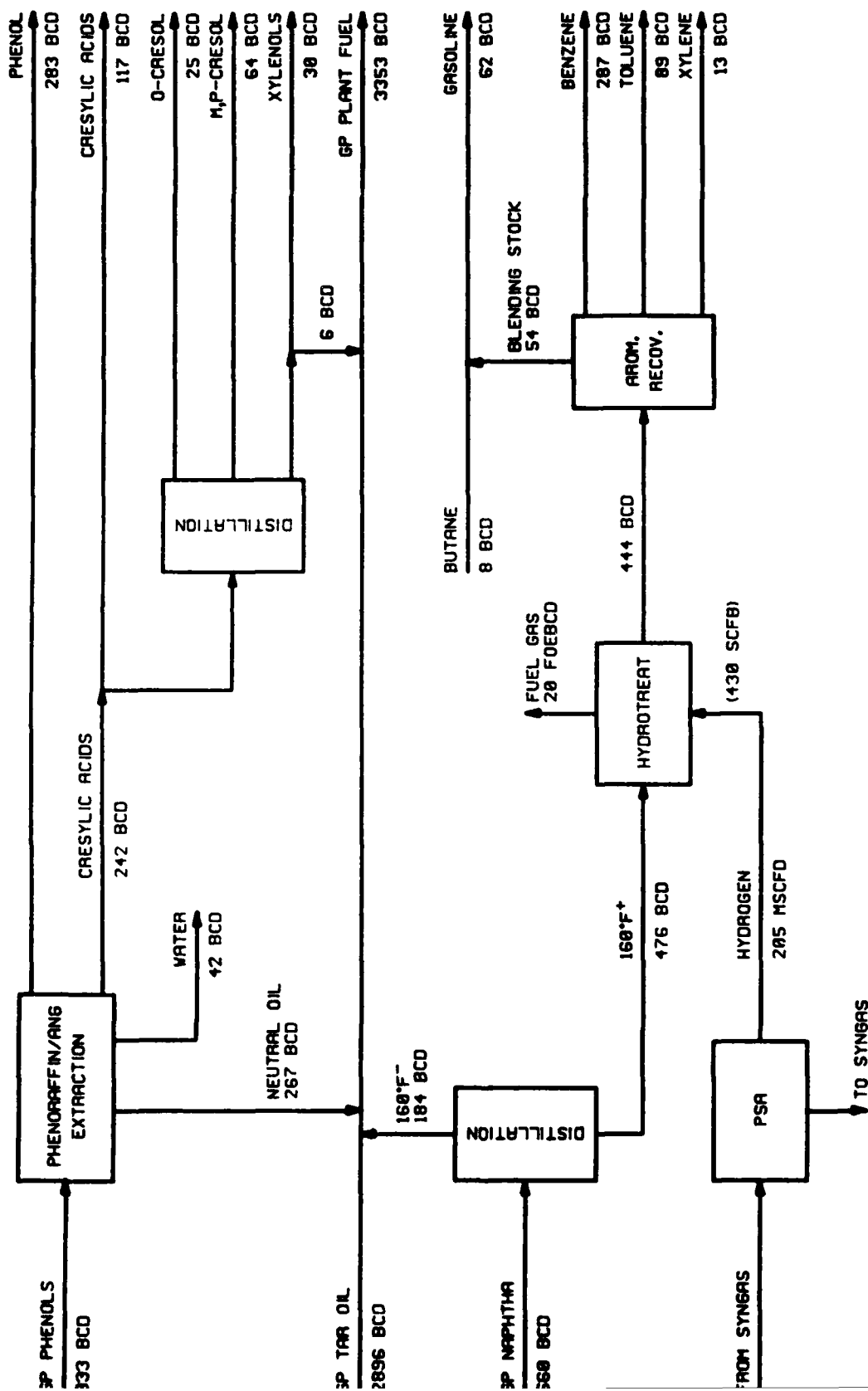


FIGURE 15
GREAT PLAINS LIQUID BY-PRODUCTS USE
CASE 7: MAXIMUM PROFIT CASE

The Great Plains phenol stream goes to a Phenoraffin/ANG extraction unit to recover purified phenol and cresylic acids. The remaining neutral oil stream is burned for plant fuel. Some of the cresylic acids are fractionated to make o-, m-, and p-cresol, and xylenols, which are available for sale up to the market limit (10 percent of U.S. market).⁽⁹⁾ If the cresylic acid fractionation system is built to handle the entire stream volume, the LP chooses to fractionate enough to meet the market limit on all stocks, even though this means dumping 6 BD excess xylenols to fuel. The arbitrary market limit is explored as a sensitivity later in this report.

Since the maximum profit case uses only the Great Plains naphtha and phenol streams, the tar oil stream is available to make jet fuels. The remaining cases use maximum available tar oil to make jet fuel. The profit in these cases is less than when no jet fuels are produced. Profitability for these cases depends on replacement fuel cost, which is reported later as a sensitivity.

Case 2 - Profitable JP-4.--The profitable JP-4 case is shown in Figure 16. The Great Plains tar oil is hydrotreated and hydrocracked as in Case 1. The Great Plains naphtha and phenol streams are processed to make chemicals and BTX, as in Case 7, but the amount of BTX extraction is adjusted to balance the octane requirements for blending with naphthas from the hydroprocessors to produce an unleaded gasoline product. The design basis for this case specifies aromatics recovery and cresylic acid fractionation units big enough to process the entire stream, since the incremental cost is small. As in the maximum profit case, a small volume of xylenols are diverted to fuel. Hydrogen for all three hydroprocessors is obtained from a PSA unit.

With the price structure assumed, it is better to sell some naphtha streams as reformer feed to existing refiners than to build a new reformer unit at the Great Plains site. Table 9 shows this sales option was selected in Cases 2, 4, and 6.

Case 4 - Profitable JP-8.--The profitable JP-8 case is shown in Figure 17. The Great Plains tar oil is hydrotreated and hydrocracked under almost identical conditions as in Case 3. The Great Plains naphtha and phenol

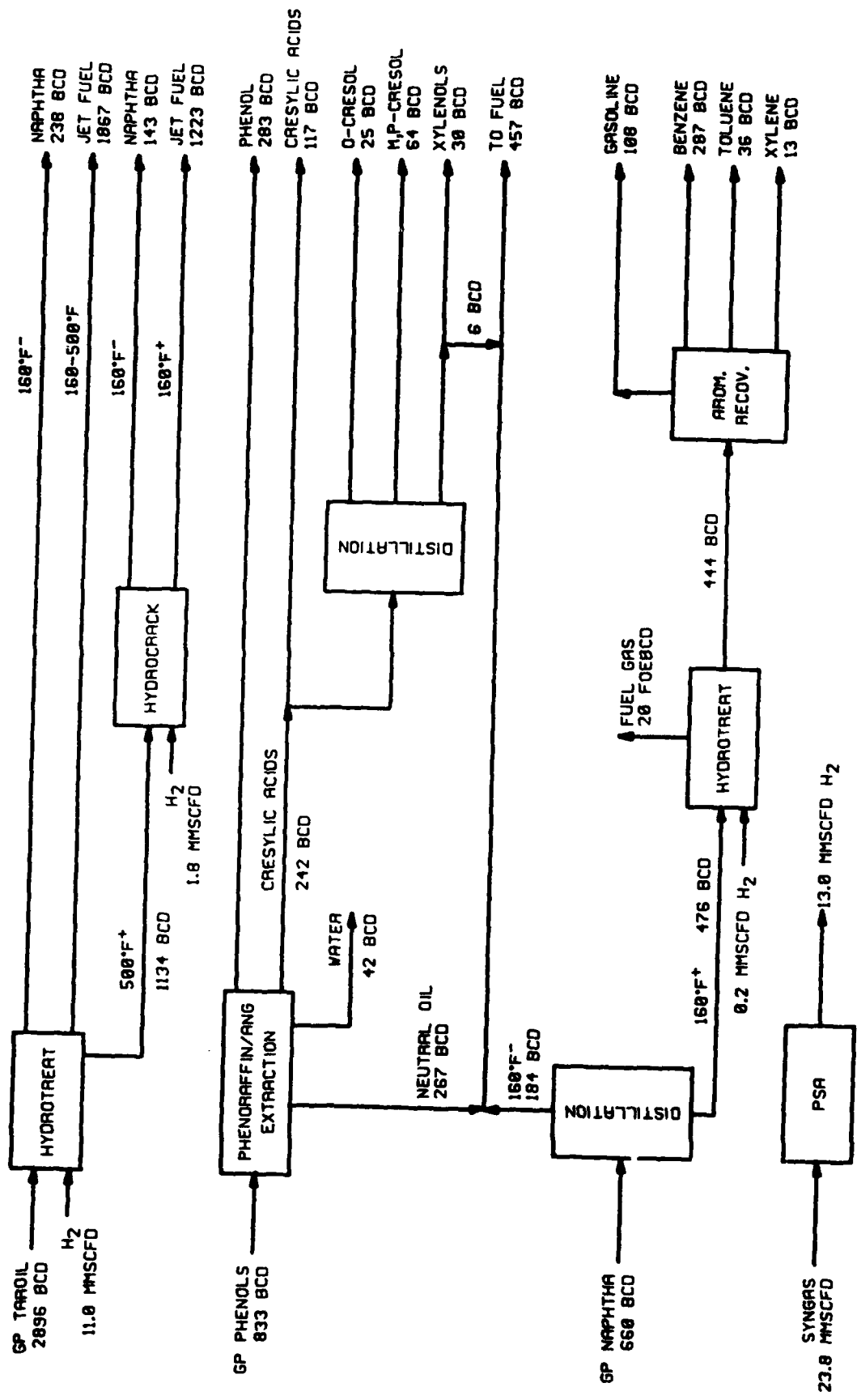


FIGURE 16
GREAT PLAINS LIQUID BY-PRODUCTS USE
CASE 2: PROFITABLE JP-4 CASE

TABLE 9

SUMMARY OF PROCESS DESIGN AND COST ESTIMATES

<u>Case</u>	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>6</u>	<u>7</u>
<u>Unit Production Rates (BPCD)*</u>						
Jet Fuel	3698	3062	2864	2297	1744	-
Reformer Feed/Gasoline	-	437	-	1190	511	41
Phenol	-	285	-	285	285	285
Cresylic Acids	-	236	-	236	236	236
BTX	-	294	-	398	398	398
 <u>Capital Cost (THSD \$)</u>						
Jet Fuel/Naphtha	51,430	43008	44,799	42978	31111	-
Phenol/Cresylic Acids	-	20124	-	20124	20124	20124
BTX	-	15000	-	17664	17664	17664
TOTAL	51,430	78,132	44,799	80,766	68,899	37,788
 <u>Operating Costs (\$/D)</u>						
Utilities (Incl. SNG Equiv.)	79,230	89,980	51,130	90,947	74,416	21,121
Cat & Chem	2400	3416	2963	3331	3140	1373
STM Usage (#/H) HP	-	54,700	-	58,200	58,200	58,200
MP	5300(Exp)	8900	6945(Exp)	9,300	9,300	15,900
LP	-	6900(Exp)	-	7050(Exp)	7050(Exp)	6900(Exp)
Personnel	40	76	40	76	68	56

*Appendices give production rates based on barrels per stream day, not barrels per calendar day.

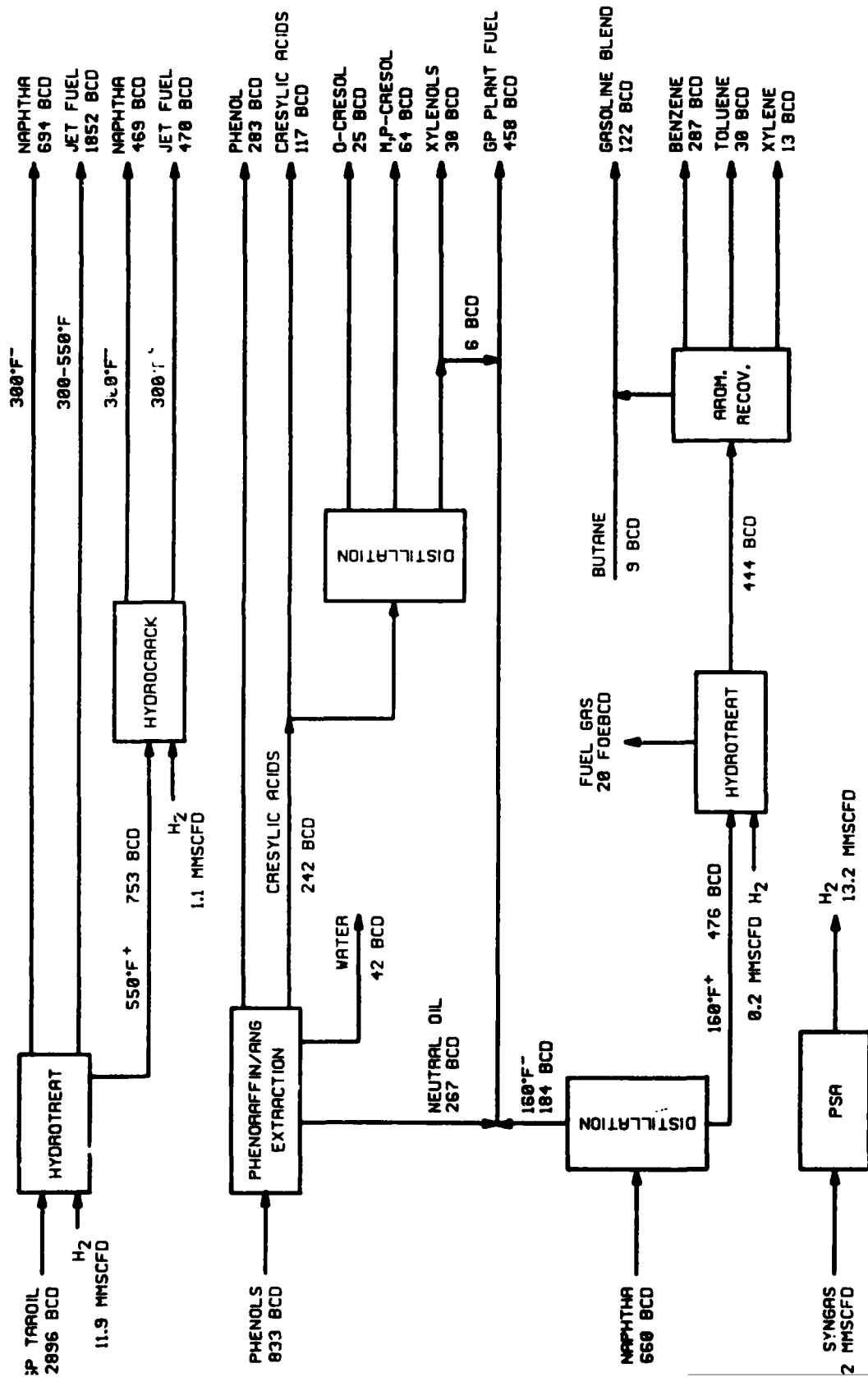


FIGURE 17
GREAT PLAINS LIQUID BY-PRODUCTS USE
CASE 4: PROFITABLE JP-8 CASE

streams are processed to make chemicals and BTX as in Case 7, but additional toluene is added to the gasoline blending stock in the aromatics recovery unit to balance the octane requirements for blending with naphthas from the hydroprocessors. The commercial design basis for this case would build the cresylic acid fractionation unit big enough to process the entire stream. Hydrogen for all three hydroprocessors is obtained from a PSA unit.

Case 6 - Profitable JP-8X.--The profitable JP-8X case is shown in Figure 18. To produce a volume of jet fuel comparable to Case 5, the LP hydrotreats the neutral oil from Phenoraffin extraction along with 85 percent of the 750°F- Great Plains tar oil. The remaining tar oil and the 750°F+ cut are used as fuel. The neutral oil stream, not available in Case 5, can be hydrotreated to produce JP-8X more cheaply than from tar oil. The Great Plains naphtha and phenol streams are processed to make chemicals and BTX, as in Case 7, but the amount of BTX extraction is adjusted to balance the octane requirements for blending with naphtha from the hydrotreater. The design basis for this case would build the aromatics recovery and cresylic acid fractionation units big enough to process the entire stream. Hydrogen for both hydroprocessors is obtained from a PSA unit.

A summary of the feeds and products for the LP Cases 1-7 are shown in Table 10, along with unit capacities for each case.

c. Process Design and Cost Estimate

Based on the LP conceptual process schemes, Lummus has developed conceptual process designs and cost estimates for the various product slates, with the exception of Case 5. Table 9 summarizes Lummus production rates, capital and operating costs, which have been developed for each case. Appendices B through H detail the results. Production rates in Tables 9 and 10 are different because Table 9 is based on a preliminary process design basis, whereas Table 10 gives rates which were re-optimized based on capital and operating costs from the preliminary process design and all the process and blending limits in the LP. Investments and utilities are also different in Tables 9 and 10 because of this re-optimization.

In agreement with Amoco and the DOE, the phenol stream process (Phenoraffin/ANG) design is based on data received from ANG Coal Gasification Co.⁽⁴⁾ Consequently, the phenol extraction and cresylic acid distillation areas are not based on proven, licensable technology and remain the least certain processing blocks in the plant.

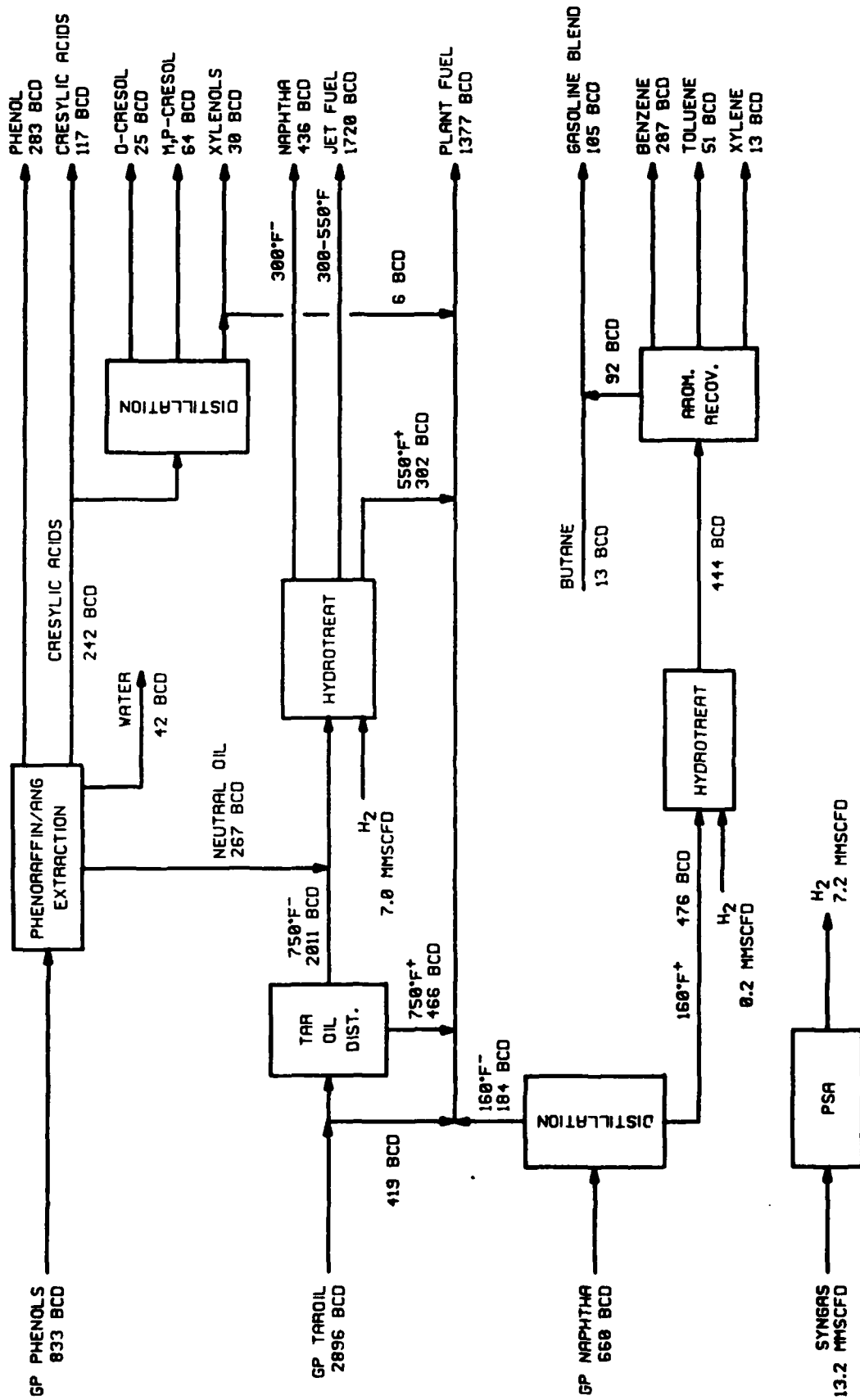


FIGURE 18
GREAT PLAINS LIQUID BY-PRODUCT USE
CASE 6: PROFITABLE JP-8X CASE

TABLE 10
GREAT PLAINS LIQUID BY-PRODUCTS CASE SUMMARY
(Total Upgrading Process)

<u>Economics</u>	<u>Max Profit</u>	<u>JP-4</u>		<u>JP-8</u>		<u>JP-8X</u>	
		<u>Max.</u>	<u>Prof.</u>	<u>Max</u>	<u>Prof.</u>	<u>Max</u>	<u>Prof.</u>
Profit, M\$/CD	40.8	-3.2	32.7	-16.2	24.8	-14.7	29.0
Profit, MM\$/yr	14.9	-1.2	11.9	-5.9	9.0	-5.3	10.6
Investment, MM\$	37.0	53.1	87.1	52.7	89.3	33.7	68.5
<u>Feedstocks, BCD</u>							
G.P. Naphtha	660	0	660	0	660	0	660
G.P. Phenol	833	833	833	0	833	0	833
G.P. Tar Oil	0	2896	2896	2896	2896	2896	2896
Syngas, MMSCFD	0.4	30.4	23.8	23.7	24.2	14.3	13.2
<u>Products, BCD</u>							
Gasoline	62	466	336	239	297	0	165
Reformer Feed	0	0	154	983	988	379	376
Jet Fuel	0	4052	3090	2323	2322	1763	1720
BTX	389	0	336	0	330	0	351
Oxygenated Chem.	519	0	519	0	519	0	519
Liq. Fuel	457	0	457	0	454	947	1377
<u>Unit Cap., BCD</u>							
Aromatics Rec.	444	0	444	0	444	0	444
Phenoraffin/ANG	833	0	833	0	833	0	833
Tar Oil Dist.	0	0	0	0	0	841	2477
Hydrocrack.	0	1159	1134	753	754	0	0
Hydrotreat.	0	3729	3563	4797	4796	2379	1925
Naphtha Dist.	660	0	660	0	660	0	660
Naphtha Hydrotrt.	476	0	476	0	476	0	476
PSA, MMSCFD	0.4	30.4	23.8	23.7	24.2	14.3	13.2

As a result of the work accomplished during this task, two important items for the economic success of the by-product processing must be resolved.

The first and most apparent is the development of a process to provide a saleable cresylic acid product slate. Product specifications and limitations on production, if any, must be developed. Also, allowable levels of guaiacol and catechol in the cresylic acids must be addressed. One means of separating these close-boiling materials from the desired products may be through the use of crystallization. Although the compounds have close boiling points, their melting points differ substantially. It may be desirable to separate the components into narrow boiling point ranges and then purify each cut by fractional crystallization. Pilot plant work should be undertaken to develop an effective means of recovering these saleable products.

Secondly, Great Plains fuel requirements will increase significantly in some cases to replace the fuel presently being provided by the by-products. Since the replacement fuel required for the various profitable cases (1-4 MBD) is comparable to the residual fuel in LPG consumption of all North Dakota (3 MBD),⁽²⁾ alternative fuel sources must be studied and incorporated into the design basis of these units.

3. Subtask 1.3 Economic Analysis

a. Assumptions and Basis

As is the case with all preliminary evaluations, a number of assumptions were made to reach the economic results which are presented in this section. Before presenting the results, a review of these assumptions is warranted.

In all seven cases,

1. It is assumed that the existing Great Plains plant can supply adequate high pressure steam and other utilities and power to the liquid by-product upgrading plant. Cost and availability information was supplied by ANG.

2. It is assumed that any H_2S generated by the upgrading facility can be processed to produce sulfur by the existing Great Plains sulfur plant. No capital or operating allowance has been made for a separate sulfur plant for by-product upgrading.
3. It is assumed that any ammonia and wastewater produced by the upgrading facility can be handled by the existing Great Plains ammonia recovery and wastewater treatment plants.
4. It is assumed that the existing boilers which currently burn the tar oil, naphtha, and phenolic streams can be modified to burn the selected alternative fuel with little or no capital investment. No capital has been included in the estimates for the various cases to modify the boilers. Furthermore, no allowance has been made for storage of the replacement fuel.
5. It is also assumed that the various processes which are considered by the LP can produce the required product at the commercial specifications for that product. As pointed out previously, this assumption is particularly important for the processes which handle the phenolic stream. Amoco has no commercial experience with these processes.
6. Numerous economic assumptions are made. It is assumed that:
 - a. The upgrading plant is 100 percent equity financed.
 - b. The upgrading facility is built in two years and started up in one year.
 - c. A 10 percent real rate of return is required on invested capital.
 - d. No land purchase is necessary.
 - e. Upgrading plant life is 20 years and the plant has zero salvage value.

- f. Federal and North Dakota state taxes combined are 40 percent of gross profit.
- g. The plant onstream factor is assumed to be 0.91 for tar oil and naphtha processing and 0.89 for phenolic processing.
- h. The upgrading plant is depreciated over a 10 year period using the Tax Recovery Act of 1986 schedule.
- i. Various levels of investment contingency (10 to 30 percent) have been added according to process definition and maturity.

These assumptions are consistent with a capital charge factor of 16 percent/year.

Note that assumptions 1 through 4 will be dealt with during Task 4.

b. Economic Results

The conceptual process schemes to upgrade the by-products were developed by the Great Plains LP, based on Amoco estimates of capital and operating costs for the various process steps. As discussed above, Lummus developed detailed process designs and capital and operating costs for each conceptual case. The LP was revised with these Lummus cost figures and the economics of each case was estimated, assuming a replacement fuel cost of \$2.15/MMBtu. In the discussion that follows, profit is above a 10 percent real rate of return on invested capital, as all costs include a capital charge on capital employed in the upgrading plant.

Table 11 shows the investment breakdown by unit. Table 12 summarizes the economics of the cases, broken down by cost component. All the maximum jet fuel cases show a loss (negative profit) because the total cost of upgrading

TABLE 11

GREAT PLAINS INVESTMENT AND UTILITIES SUMMARY

<u>MM\$</u>	<u>Max Profit</u>	<u>JP-4</u>		<u>JP-8</u>		<u>JP-8X</u>	
		<u>Max.</u>	<u>Prof.</u>	<u>Max</u>	<u>Prof.</u>	<u>Max</u>	<u>Prof.</u>
Aromatics Rec.	12.3	0	12.3	0	12.3	0	12.3
Phenoraffin	19.4	0	19.4	0	19.4	0	19.4
Tar Oil Dist.	0	0	0	0	0	3.5	4.3
Hydrocrack.	0	13.2	13.0	11.2	11.2	0	0
Hydrotreat.	0	28.1	27.5	31.3	31.3	23.2	21.1
Naphtha Dist.	0.2	0	0.2	0	0.2	0	0.2
Naphtha Hydro.	4.5	0	4.5	0	4.5	0	4.5
PSA	0.5	10.5	8.8	8.8	9.0	6.2	5.9
Power Dist.	0.1	1.4	1.3	1.4	1.4	0.8	0.8
Total	<u>37.0</u>	<u>53.1</u>	<u>87.1</u>	<u>52.7</u>	<u>89.3</u>	<u>33.7</u>	<u>68.5</u>

Utilities⁽¹⁾

Cat & Chem, \$/D	542	3533	3920	4470	5014	4389	4092
Fuel, FOEB/D	1068	4071	4330	3094	4336	2159	2935
Power, MW	0.2	6.2	5.8	6.0	6.3	3.2	2.9
Cool Wat., M gpm	2.9	2.2	4.9	2.3	5.2	1.0	3.8
Proc Wat., gpm	3	31	33	38	40	17	16

Note: Steam costs or credits are allocated in this table to fuel and cat & chem.

(1) These utilities do not agree with those in Table 9 because of reoptimization by the LP.

TABLE 12

GREAT PLAINS ECONOMICS SUMMARY

Cash Flow M\$/CD	Max Profit	JP-4		JP-8		JP-8X	
		Max.	Prof.	Max	Prof.	Max	Prof.
Net Sales (1)	79.9	104.3	163.2	78.4	158.5	47.9	127.7
Fuel (2)	-14.6	-65.4	-66.7	-52.0	-66.9	-34.1	-48.3
Cat and Chem	-0.5	-3.5	-3.9	-4.5	-5.0	-4.4	-4.1
Utilities (3)	-0.9	-6.5	-6.5	-6.4	-7.3	-3.3	-3.7
MTIO (4)	-4.1	-5.8	-9.5	-5.8	-9.8	-3.7	-7.5
Fixed Costs (5)	-2.6	-2.8	-5.3	-2.7	-5.2	-2.3	-4.9
Capital Recov.(6)	-16.4	-23.5	-38.5	-23.3	-39.5	-14.9	-30.3
Total Profit	40.8	-3.2	32.7	-16.2	24.8	-14.7	29.0
Total MM\$/yr	14.9	-1.2	11.9	-5.9	9.0	-5.3	10.6

Notes:

- (1) Includes naphtha, gasoline, BTX, and chemicals, less the cost of purchased gasoline blending stocks (e.g., butane).
 (2) Includes Great Plains naphtha, tar oil, phenol, and hydrogen removed from syngas, as well as purchased fuel, less credit for fuel returned to the Great Plains pool. Hydrogen is priced at a premium over fuel value. Replacement fuel is assumed to be LPG at \$2.15/MMBtu.
 (3) Includes power, process water, and cooling water. Steam costs and credits are allocated to fuel and catalysts and chemicals.
 (4) Maintenance, taxes, insurance, and overhead charges.
 (5) Primarily operating labor.
 (6) Ten percent real rate of return, 5 percent inflation, 2-year construction, 1-year startup.

Jet Fuel, BCD Production Rate	0	4052	3090	2323	2322	1763	1720
Jet Fuel Unit Profit/Loss to Breakeven, \$/B	-	-0.8	10.6	-7.0	10.7	-8.4	16.9

the feeds to jet fuels and of replacement fuel exceed the revenue generated by jet fuel sales. Table 6 shows that oxygenated chemicals and BTX are valued high enough above feed prices that cases which make these products from the Great Plains phenol and naphtha streams generate a profit. The maximum profit case could potentially add about 15 MM \$/yr to the Great Plains annual profits. Profitability of the cases which make jet fuels, besides chemicals and BTX, are between 9 and 12 MM \$/yr.

Another way of looking at the economics is to calculate how much the jet fuel price would have to be subsidized to break even for the maximum jet fuels cases, and how much the jet fuel price could drop to break even for the profitable jet fuels cases. These values, shown in Table 12, suggest that a subsidy less than a \$9/B is needed to break even for the maximum jet fuels cases; in fact, the JP-4 case subsidy is less than \$1/B. If BTX and chemicals by-products are produced, jet fuel sales prices could be lowered by \$10-17/B and still break even.

c. Sensitivities

The economics discussed above are affected by several key assumptions. The effects of altering these assumptions is explored in this section. The objective of this section is to provide a scoping estimate of the impact of these assumptions on the economics. Lummus has not confirmed Amoco's estimated capital and operating costs for some of the process schemes discussed in this section.

The economics presented above assume that the Great Plains stocks could be replaced with other fuel stocks at \$2.15/MMBtu. This value is roughly the current cost of LPG in the North Dakota area. Table 13 shows the values of LP streams if other fuel costs are assumed from \$1.50/MMBtu to \$5.00/MMBtu, by adjusting for the Btu content of the various streams. The LP has been optimized with these values, and the effect of fuel price on profitability for the maximum jet fuels cases is shown in Figure 19. Higher fuel costs have a strong negative effect on profit. Fuel costs at least below \$2.00/MMBtu are

TABLE 13
EFFECT OF FUEL PRICES ON STREAM VALUES

Stream	Unit	Fuel Price, \$/ MM Btu			
		1.50	2.15	3.80	5.00
Natural Gas	FOE B	9.45	13.57	23.94	31.50
LPG	Bbl	5.26	7.57	13.32	17.53
n-Butane	Bbl	11.76*	11.76*	15.16	19.95
Grt P1 Naphtha	Bbl	8.85	12.69	22.42	29.50
Grt P1 Phenol	Bbl	7.28	10.43	18.43	24.25
Grt P1 Tar Oil	Bbl	9.12	13.07	23.10	30.40
Syngas for H2	MSCF	0.67*	0.67*	1.08	1.42
Utilities Fuel	MMBtu	1.50	2.15	3.80	5.00

* Current price is above fuel value. Current price is used.

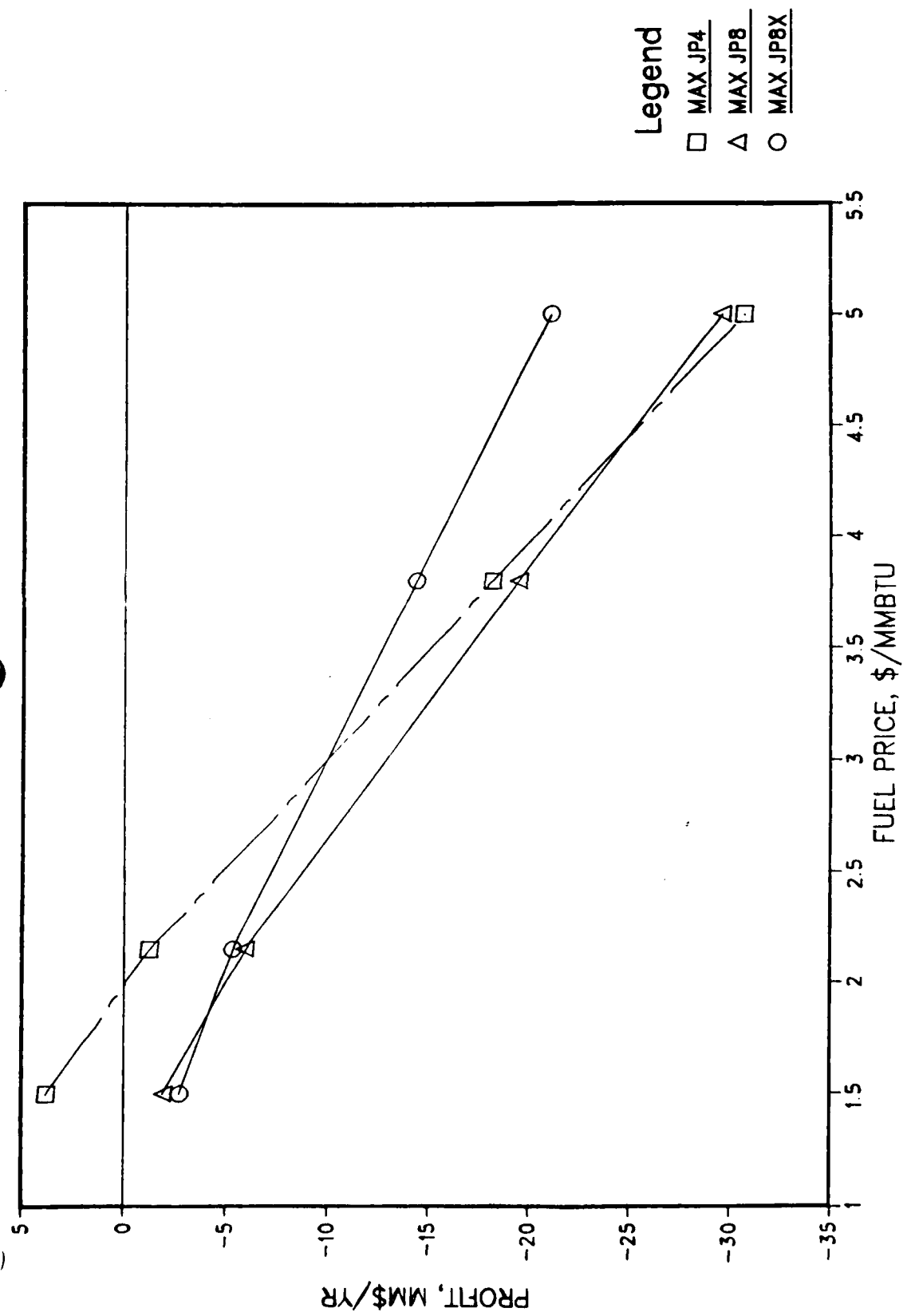


FIGURE 19
MAXIMUM JET FUEL CASES
EFFECT OF FUEL PRICE

needed to generate a profit in the maximum JP-4 case. Fuel costs below \$1.50/MMBtu would be required to generate a profit in the JP-8 and JP-8X cases.

Figure 20 shows the effect of varying product slates along with fuel price. The maximum profit case retains profitability for all fuel values studied, although profit falls by about \$2 MM/yr for each \$1/MMBtu increase in fuel costs. The maximum JP-4 case is copied from Figure 19. If the phenolic stream in the maximum JP-4 case is processed to make oxygenated chemicals while processing the tar oil to make JP-4, the profitability is increased about \$10 MM/yr. This processing is profitable if fuel costs are below about \$3.10/MMBtu. If, in addition, the Great Plains naphtha stream is processed for BTX, profits increase further at low fuel costs. Break even fuel cost is about \$3.20/MMBtu for this processing.

Figure 21 shows the effect of fuel cost on profitability for the profitable jet fuels cases. The JP-8 case is profitable for all fuel costs below \$3.00/MMBtu; the JP-4 case is profitable below \$3.20/MMBtu; and the JP-8X case is profitable below \$3.50/MMBtu. Profits are below the case which produces no jet fuels, and these cases are also more sensitive to fuel costs.

d. Hydrogen Cost

Table 14 shows that the hydrogen cost component in the maximum profit case is negligible, so that case would not be affected by hydrogen cost variation. The cases which produce jet fuels have hydrogen costs between 3 and 7 MM\$/yr. Doubling hydrogen costs would therefore decrease profits by 3 to 7 MM\$/yr.

e. Effect of Oxygenated Chemical Market Size

Figure 22 shows the effect of imposing various limits on the sales of cresols, xylenols, and cresylic acids as a percent of the U.S. market.⁽⁹⁾ The starting basis for this discussion is Case 7 at 10 percent of the U.S. market for these oxygenated chemicals. As Figure 15 shows, in Case 7 the Great

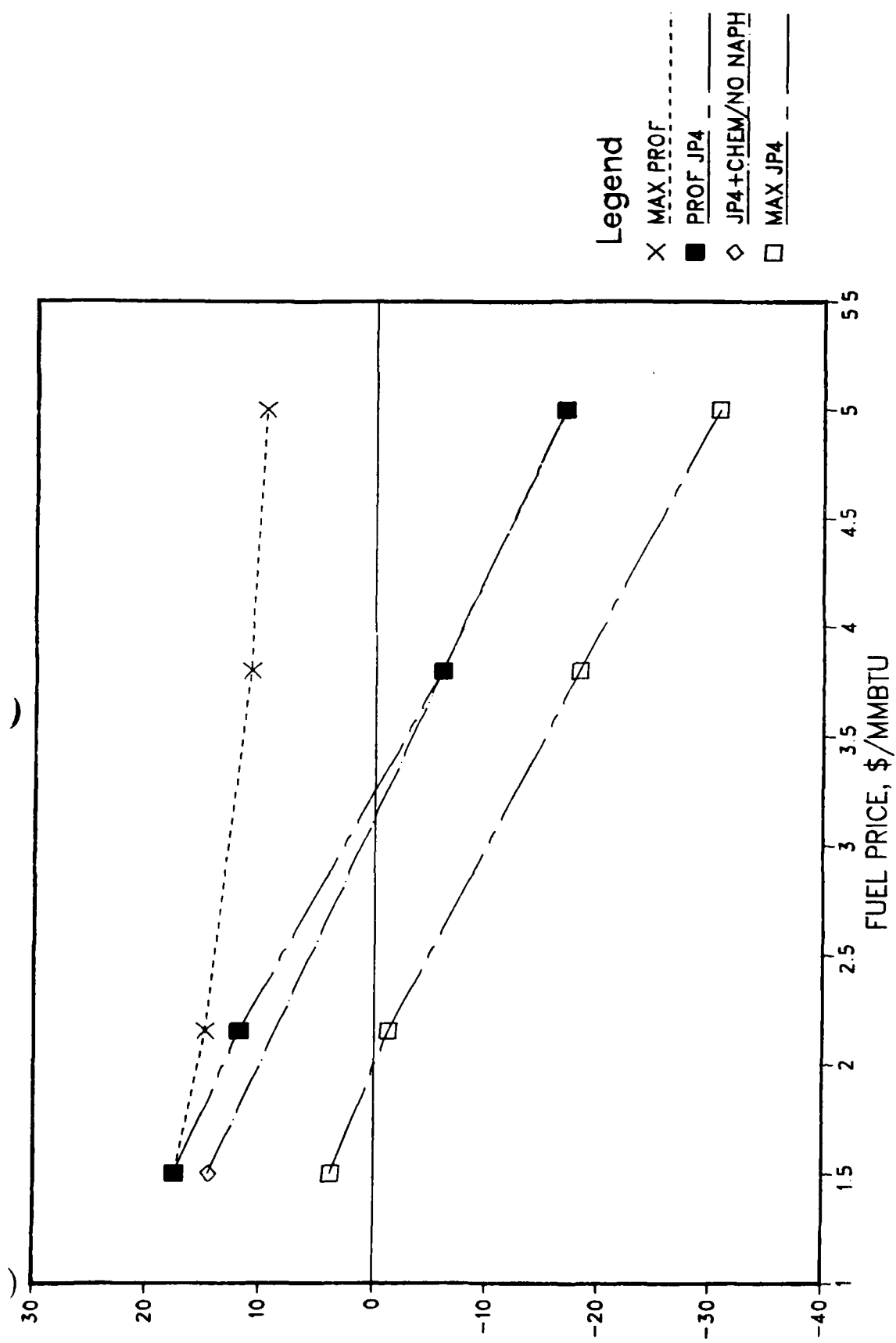


FIGURE 20
CASES WITH JP-4
EFFECT OF FUEL PRICE

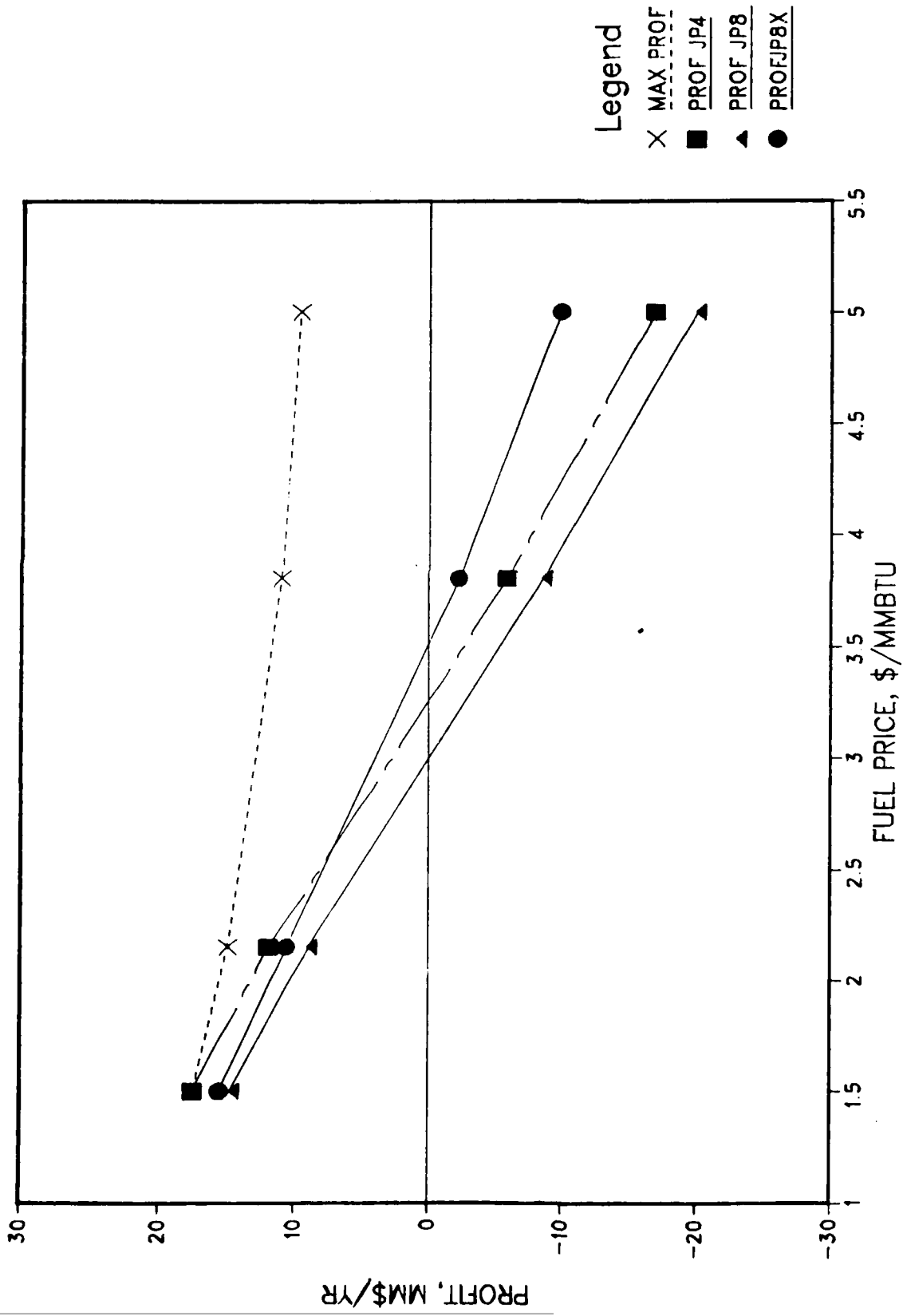


FIGURE 21
 PROFITABLE CASES
 EFFECT OF FUEL PRICE

TABLE 14
EFFECT OF HYDROGEN COST ON PROFITABILITY

<u>Case</u>	<u>Syngas, MMSCFD</u>	<u>Cost*</u>	
		<u>M\$/CD</u>	<u>MM\$/yr</u>
Maximum Profit	0.4	0.2	0.1
Maximum JP-4	30.4	20.4	7.4
Maximum JP-8	23.7	15.9	5.8
Maximum JP-8X	14.3	9.6	3.5
Profitable JP-4	23.8	15.9	5.8
Profitable JP-8	24.2	16.2	5.9
Profitable JP-8X	15.2	8.8	3.2

* at \$1.23/MSCF of hydrogen removed.

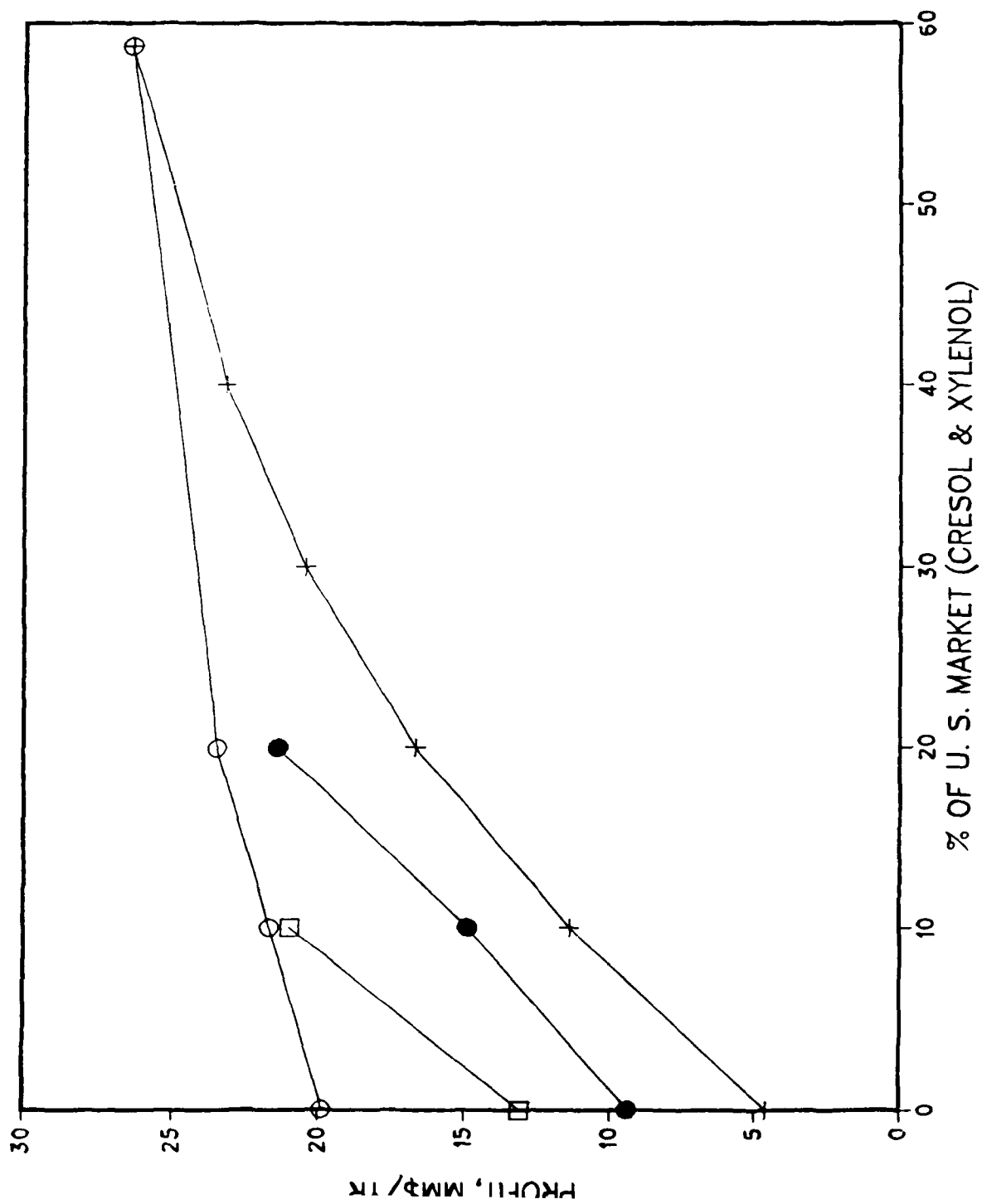


FIGURE 22
MAXIMUM PROFIT CASE
EFFECT OF CHEMICALS SALE

Plains phenols stream is extracted to make phenol and cresylic acids. Then some cresylic acids are fractionated to make maximum amounts of cresols and xlenol. Remaining cresylic acids are sold.

If cresylic acids sales are limited to lower levels, a Dynaphen unit is built to convert any non-fractionated cresylic acids into benzene and phenol. Fractionation for sales as cresol and xlenol is always favored over cresylic acid sales or Dynaphen processing. Figure 23 shows the process units when Dynaphen is maximally used (at 0 percent of cresol, xlenol, and cresylic acid sales). In this process scheme, all cresylic acids from extraction are processed in Dynaphen. It may be possible to process the Great Plains phenol stream directly with Dynaphen without prior extraction by distilling phenol from the cresylic acid. This would improve the economics of this case by about \$4 MM/yr in capital and operating costs. All points in the lower left of Figure 22 involve some Dynaphen processing, as illustrated by Figure 24.

As sales volumes are increased from 10 percent of the U.S. market, cresol and xlenol sales are satisfied by fractionating additional cresylic acids. When cresol and xlenol and cresylic acid volumes combined reach or exceed 30 percent of the U. S. market, the tar oil stream is fractionated so that additional chemicals can be extracted from 450°F- tar oil. Ultimately, all 450°F- tar oil is extracted, and all cresylic acids from both streams are fractionated into cresols and xlenols. Figure 22 shows that this is reached at about 60 percent of the U.S. xlenols market, although cresols volume are feedstock limited to 25 to 35 percent of the U.S. market. The change of slope in the lines observed in Figure 22 at market size greater than 20 percent of U.S. market is due to the fact that the 450°F- tar oil is leaner in oxygenated chemicals than the phenolic stream. Also shown in Figure 24 are the estimated regions of Dynaphen and 450°F- tar oil processing.

These economics assume the sales prices listed in Table 6. It is doubtful if these prices would exist if Great Plains significantly expanded the U. S. supply without further market development.

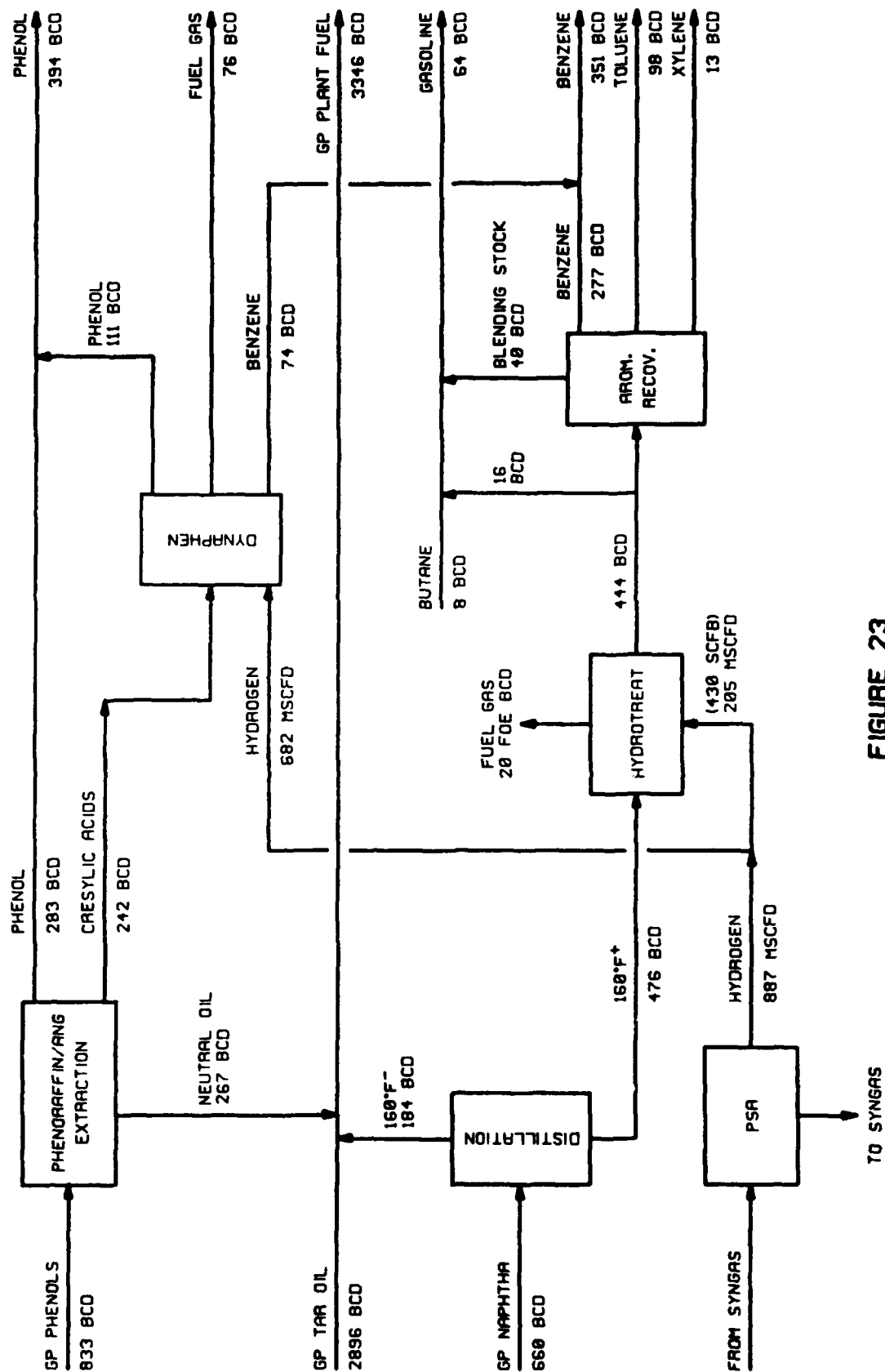


FIGURE 23
GREAT PLAINS LIQUID BY-PRODUCTS USE
WITH NO CRESEOLS OR CRESYLIC ACIDS

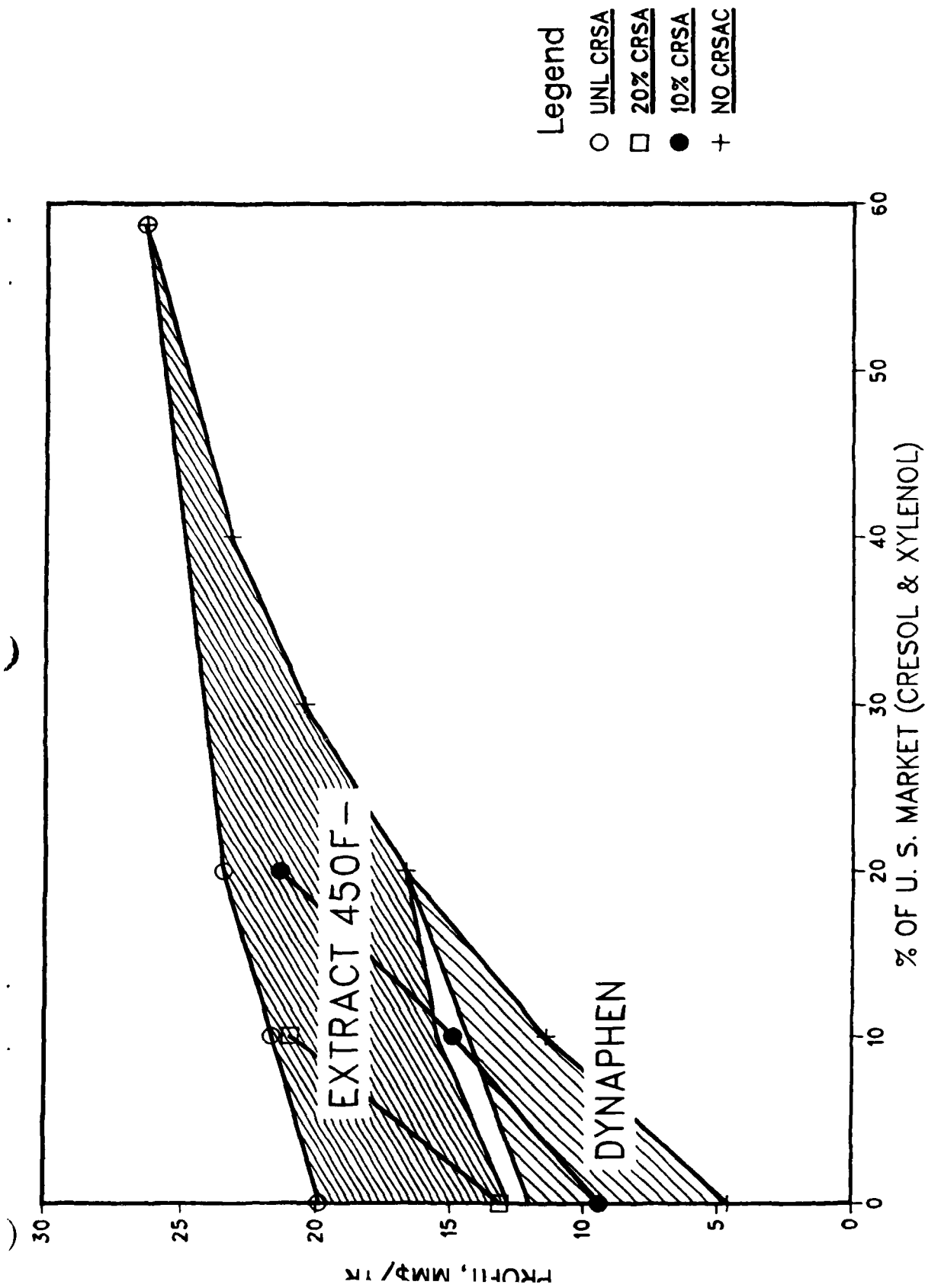


FIGURE 24
MAXIMUM PROFIT CASE
CHEM PROCESS ALTERNATIVES

SECTION V

SUMMARY AND CONCLUSIONS

A preliminary evaluation of the upgrading alternatives for the by-product streams (naphtha, phenols, and tar oil) from the Great Plains Gasification plant has been completed. The evaluation is based on (1) Amoco and other laboratory analytical analyses of the three streams, (2) the Western Research Institute scoping studies on hydrotreating the tar oil, (3) a market analysis of the various products by J. E. Sinor Consultants, Inc., (4) Amoco's proprietary process models for various refining technologies, (5) Lummus's cost estimates for the various required processes and (6) literature and ANG information on processes to upgrade and separate the phenolics stream. As a result of the analyses, seven possible upgrading schemes for the by-product streams were developed. The product slates in these schemes ranged from the maximum production of the various grades of jet fuel (JP-4, JP-8, JP-8X) to a slate which produced various chemicals (cresylic acid, phenol, cresols, xylenol, benzene, toluene, and xylene). As a result of this preliminary evaluation, the following conclusions have been reached:

1. The aging of the various by-product streams under storage conditions is minimal. Thus the quality of the by-product streams will not deteriorate in intermediate storage.
2. The quality of the by-product streams produced from the gasifier does not vary greatly with time. Thus the by-product quality bases used in our conceptual designs are representative.
3. The various grades of jet fuel can be produced from the tar oil at Great Plains, but not economically.
4. The phenolic and naphtha streams have the potential to significantly increase revenues at Great Plains.
5. The phenolic and naphtha streams could provide sufficient revenue to subsidize the production of jet fuel.

6. The economics of jet fuel production are sensitive to replacement fuel cost.
7. The amount of cresols and cresylic acid which can be marketed is a concern.
8. Processing the phenolic stream to produce cresols and cresylic acid should be demonstrated. The potential problems with contaminants such as guaiacol, catechol, and other contaminants should be assessed.

SECTION VI

RECOMMENDATIONS

An important goal of Task 1 is to recommend experiments which would be carried out in Task 2 to confirm the major (significant) assumptions of Task 1. Recommended experiments are:

For the tar oil by-product:

1. Additional hydrotreating data are needed to confirm yields, upgraded tar oil qualities, process conditions, and hydrogen consumption. These variables have the greatest impact on the economics.
2. Heat release as a function of product quality should be estimated to confirm the choice of an expanded bed over a fixed bed hydrotreater.
3. The impact on hydrotreating requirements of blending the phenolic stream with the tar oil stream to maximize JP-4 production should be determined.
4. The effect of catalyst type on JP-8X properties should be determined. (Can a catalyst be found which will allow saturation of sufficient aromatic rings, without opening the rings, so that the JP-8X specifications on aromatics and paraffin content can be met?).
5. The 500°F+ and the 550°F+ fractions of the hydrotreated tar oil should be hydrocracked to determine yield, product quality, process conditions, hydrogen consumption and heat release.

For the naphtha stream:

- The naphtha stream should be hydrotreated to determine yields and product quality (mainly heteroatom content of hydrotreated naphtha).

For the phenolic stream:

--The Phenoraffin or equivalent process should be demonstrated with the Great Plains phenolics stream. The yields and qualities (purity) of the various chemicals produced should be determined. Since this is potentially the most profitable stream to process, it is important that DOE or ANG do this. Evaluation of these processes is currently beyond Amoco's experimental capabilities.

REFERENCES

1. Smith, E. B., "Jet Fuels from Coal; Third Quarterly Report," U.S. Department of Energy Research Contract No. DE-FC21-86MC11076, May 1987.
2. Sinor, J. E., "Production of Jet Fuel from Coal-Derived Liquids, Vol 1, Market Assessment for Liquid By-Products from the Great Plains Gasification Plant," AFWAL-TR-87-2042, August 1987.
3. Wilson, W. G., C. L. Knudson, and J. R. Rindt, "Feasibility of Producing Jet Fuel from GPGP By-Products; Interim Quarterly Technical Progress Report for February 17, 1987 to May 16, 1987," U.S. Department of Energy Research Contract No. DE-AC22-87PC90016.
4. Pollock, D. C., ANG letter to K. Torster (Lummus Crest), ANG letter 9440-DHD-87-041, December 30, 1987.
5. Huiber, D. T. A., and J. L. Gendler, "The Dynaphen Process: Phenol and Benzene via Thermal Hydroalkylation of Coal Liquids," Paper 3D, AIChE Annual Meeting, New Orleans, Louisiana, Nov. 8-12, 1981.
6. Sullivan, R. F., and H. A. Frumpkin, "Refining and Upgrading of Synfuels from Coal and Oil Shales by Advanced Catalytic Processes; Third Interim Report: Processing of SRC-II Syncrude," Report FE-2315-47, March 1980.
7. Ma. J. J. L., "Coal Tar Chemicals," SRI Report No. 101, December 1976.
8. Daley, D. P., letter to L. Lorenzi, DPD-87-863, Contract DE-AC22-87PC79338, October 20, 1987.
9. Anonymous, "Key Chemicals-Phenolics," Chemical and Engineering News, Nov. 4, 1985, p 12.

APPENDIX A

SATURATION KINETICS

APPENDIX A
SATURATION KINETICS

The tar oil, which at 3,180 b/sd is the largest volume liquid stream produced at Great Plains, has the greatest potential as a feedstock for jet fuel production. The very low H/C ratio and API gravity of this stream reveal that it is highly aromatic. Indeed, GC/MS analyses of the tar oil by Wilson and co-workers at the University of North Dakota Energy Research Center (UNDERC) have shown that it contains 90-95% aromatics, with the rest being paraffins.⁽³⁾ Hydrogenation data from WRI^(2,4) indicate that significant amounts of specification jet fuels can be made from the tar oil, but at the cost of very severe processing conditions ($\geq 2,000$ psig total pressure) and high hydrogen consumption (3,000-4,000 SCF/bbl). Much of this hydrogen is consumed in order to meet the jet fuel specification for aromatics content ($<25\%$).

Aromatics Saturation Kinetics

As indicated by the above discussion, a central unit for processing the tar oil and crude phenols to make jet fuels would be a hydrotreater. The WRI data indicate that the limiting factor in this hydrotreater would be saturation of the tar oil aromatics. Thus, the kinetics of tar oil aromatics saturation play an important role in the design of this hydrotreater.

The WRI hydrogenation data with the whole tar oil⁽⁴⁾ and Chevron hydrogenation data with SRC-II liquids^(5,6) were used to develop a rate expression for tar oil aromatics saturation. The SRC-II data were chosen because these liquids resemble the Great Plains tar oil in terms of heteroatom content, aromatics content, and, to a lesser extent, boiling range. The rate expression resulting from this analysis is:

$$-r_A = k_0 \exp(-E_A/RT) (P_{H_2})^{1.45} (1-X_A)^{1/2}$$

where:

- X_A = fractional conversion of aromatics
- R = gas constant
- r_A = rate of aromatics saturation, hr^{-1}
- k_0 = rate constant = $f(\text{feed, catalyst})$
- E_A = activation energy = 23,700 Btu/lb-mol
- T = average reactor temperature, $^{\circ}\text{R}$
- P_{H_2} = average reactor hydrogen pressure, psia

This expression indicates that the rate of aromatics saturation is half-order with respect to the concentration of aromatics.

APPENDIX B

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 1
MAXIMUM JP-4 PRODUCTION
SUBTASKS 1.2 & 1.3
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
JAN 30, 1988

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1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to maximize the production of type JP-4 aviation turbine fuel from Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- . Two by product streams are combined and charged to the hydrotreater (Area 100).
 - Tar Oil 47620 #/hr (3182 RPSD)
 - Phenol 14213 #/hr (920 BPSD)
- . The hydrotreater is a 3 stage expanded bed type process which removes 99% + of the sulfur, nitrogen, and oxygen compounds and begins the conversion of 500°F+ material. The hydrotreater adds a large quantity of hydrogen to the feed (3850 SCF/bbl) which results in a high heat of reaction. An expanded bed type reactor was chosen to both control and utilize the heat of reaction. Three stages were used to both control the temperature rise as well as to obtain the high efficiency associated with staging a back-mixed reactor.
- . The hydrotreater produces 6 streams:
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H₂ and CH₄.
 - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.
 - Unstabilized naphtha which is sent to the combined naphtha stabilizer in the hydrocracker (area 200). After stabilization, to control vapor pressure, the naphtha is sent to the main boiler in the SNG plant.
 - JP4 turbine fuel which is combined with JP4 produced in the hydrocracker (area 200) and sent to storage.
 - 500°F+ unconverted bottoms product which is sent to the hydrocracker (area 200).
 - Wastewater containing, NH₄OH and NH₄HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H₂S and NH₃.

- Approximately 750 #/day of spent catalyst which is shipped to a catalyst reclaimer in the same drums that the catalyst is received in.
- . The 500⁰F+ unconverted stream from the expanded bed hydrotreater (area 100) is charged to the fixed bed hydrocracker (area 200). The hydrocracker converts this material to naphtha and JP-4 turbine fuel. For this service a 5 stage unit was chosen with 65% conversion per pass. This unit also includes a naphtha stabilizer which stabilizes both the naphtha produced in the hydrotreater and hydrocracker.
- . The hydrocracker produces 4 streams in addition to JP-4.
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit of the SNG plant for recovery of the H₂ and CH₄.
 - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.
 - Stabilized naphtha which is sent to the main boiler in the SNG plant.
 - A small sour water stream which is sent to the PHOSAM unit in the SNG plant or alternatively used as part of the injection water to the hydrotreating plant.
- . Hydrogen make-up for both the Hydrotreater and the Hydrocracker is supplied from a PSA Hydrogen Unit. High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining purge gas is available at low pressure (5 psig) which has a fuel value of about 565 BTU/ft³. This H₂, CO & CH₄ rich gas is recompressed into the methanation unit of the SNG plant.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. Detailed material balances for each unit can be found in appendixes A&B. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas and naphtha produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

4109 BPSD of Tar Oil and Phenol Feed
3372 BPSD of #6 Fuel Oil
13.34 MMSCFD equivalent SNG product loss due to the syn gas feed
to the PSA unit.

Products

4278 BPSD of JP-4 turbine fuel
8.53 MMSCFD Equivalent SNG product credit due to HDT, HDC & PSA
purge gas reinjection into SNG Plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	3372 BPSD
SNG Equivalent	
of Syn Gas & Purge Gas	4.81 MM SCFD
Power	7100 kW
Cooling Water	2400 GPM (30 ⁰ F rise)
Process Water	18.5 GPM

In addition the process exports 5280 #/hr of 100 psig saturated
steam which was credited against boiler requirements.

Figure 1: Case 1 - Maximum JP-4 Production

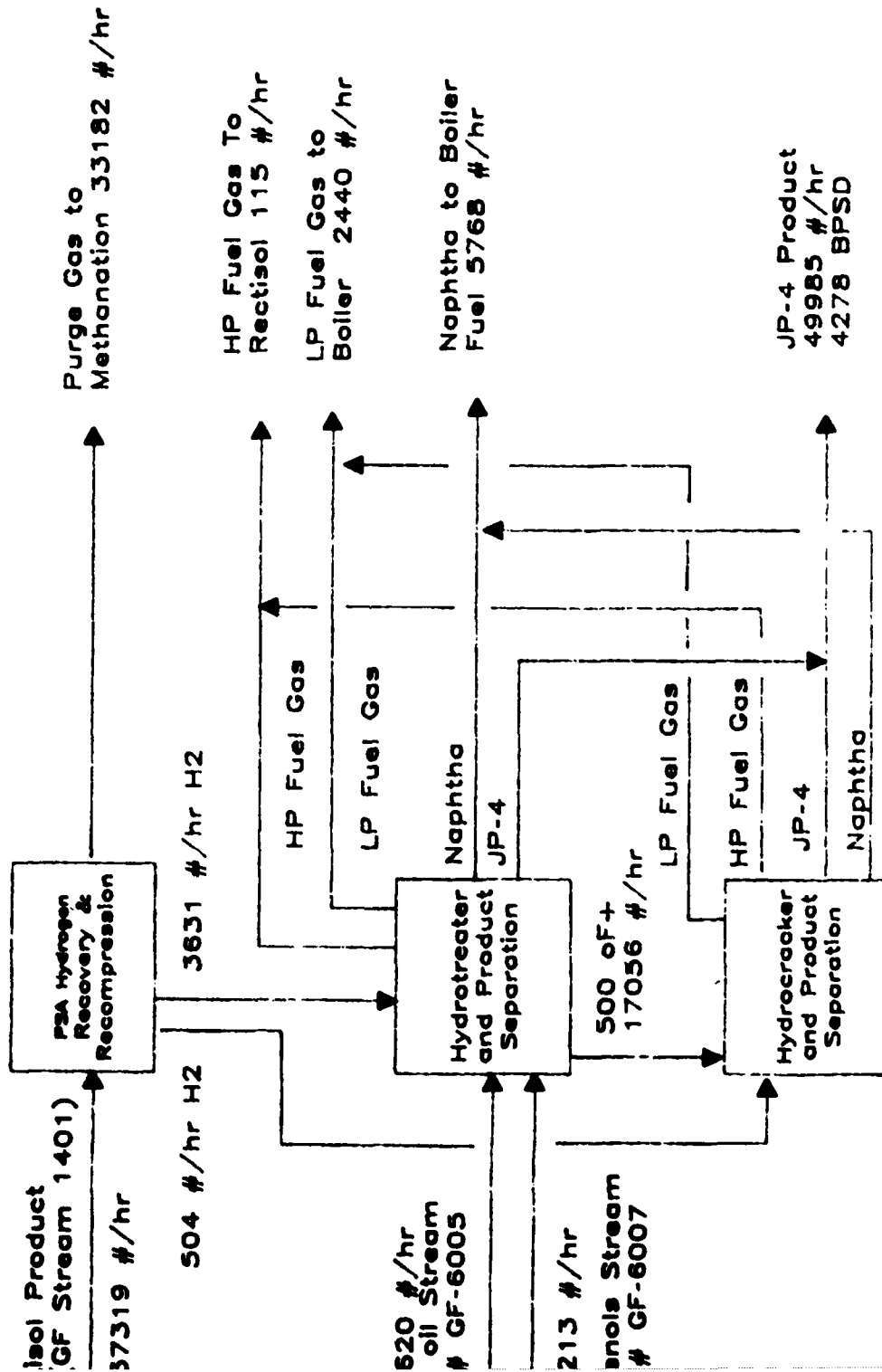


Table 1.1 Great Plains Case 1: Maximum JF4 Production

Tar Oil Feed====>	47620	#/hr	3182	BPSP
Phenol Feed====>	14213	#/hr	920	BPSP
Total Oil Feed====>	61833	#/hr	4102	BPSP
JF-4 Product====>	49985	#/hr	4278	BPSP
SNP Product Loss=>	8268	#/hr	4.8	MMSCFD
Fuel Oil Makeup==>	46646	#/hr	3372	BPSP

Expanded Bed Hydrotreater

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSP
Feeds					
H2	5.87		3631	1801.1	
Oil	100.00	1.0342	61833		4102
Total	105.87		65464		
Products					
Purge Gas	0.13		79	24.5	
Fuel Gas	2.13		1319	65.0	
Naphtha	5.50	0.7238	3403		323
JP-4	58.14	0.8208	35949		3005
500 oil	27.58	0.9465	17054		1236
H2S in SW	11.06		6839	379.9	
H2S in SW	0.35		218	6.4	
NH3 in SW	0.97		602	35.4	
Total	105.87		65464		4564

Fixed Bed Hydrocracker

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSP
Feeds					
H2	2.95		504	250.0	
500oil+	100.00	0.9465	17056		1236
Total	102.95		17560		
Products					
Purge Gas	0.21		36	10.3	
Fuel Gas	3.95		673	24.0	
Naphtha	16.49	0.6787	2813		284
JP-4	82.30	0.7563	14037		1273
H2S in SW	0.003		0.5	0.02	
NH3 in SW	0.003		0.5	0.03	
Total	102.96		17560		1558

Naphtha Stabilizer

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSP
HDT Nap	54.75	0.7238	3403		323
HCR Nap	45.25	0.6787	2813		284

PSA Hydrogen Recovery Unit (86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	8.63	5.55	903.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	2385.0	702.7	55.8	612.0	11.9	7.2	3774.6
Prod. H2	2051.1	0.2	0.0	0.0	0.0	0.0	2051.3
Purge Gas	333.9	702.5	55.8	612.0	11.9	7.2	1723.3
#/hr							
Feed Gas	4803	19681	2458	9817	357	229	37752
Prod. H2	4135	5	0	0	0	0	4140
Purge Gas	673	19676	2458	9819	357	229	33212

Fuel Gas Generated in Hydrotreating and Hydrocracking

Component	#/hr	#Mol/hr	MMBTU/hr
HDTR FG Produced	1319	65.0	25.1
HCR FG Produced	673	24.0	12.8
Stab FG Produced	448	9.8	8.2
Total Fuel Gas	2440	98.9	46.0

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	673	333.9	324	41.0
CO	19676	702.5	321	85.5
CO2	2458	55.8	0	0.0
C1	9819	612.0	1010	234.3
C2	357	11.9	1769	8.0
N2+Ar	229	7.2	0	0.0
Total	33212	1723.3	565	368.7

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft ³	BPSD
Crude Phenol	-14213	13070	-185.8			-920
Tar Oil	-47620	17000	-809.5			-3182
Fuel Gas	2440	18853	46.0	0.9	1228	
Naphtha	5768	20040	115.6			574
Export Steam	5280	1000	5.3			
Fuel Oil to Boiler	46024	18000	828.4			3317
Total	-2321		0.0	0.9		-201
Fuel Oil to Process Heaters	622	18000	11.2			45

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	13.34	90591
SNG Credit for PSA Purge	8.40	41360
SNG Credit for H ₂ purge gas	0.13	836
Total SNG Production Loss	4.81	

Water Balance Hydrotreater

Component	H ₂ O	H ₂ S	NH ₃	NH ₄ HS	NH ₄ OH	Total
Reaction Gases	6839	218	602			7659
Reaction Solution	6317			327	1015	7659
Stripping Steam	1856					1856
Softened Water	6826					6826
HDT Sour Water	14999			327	1015	16340

Water Balance Hydrocracker

Component	H ₂ O	H ₂ S	NH ₃	NH ₄ HS	NH ₄ OH	Total
Reaction Gases		0.5	0.5			1.0
Reaction Solution				0.8	0.5	1.3
Stripping Steam	556.9					556.9
HCR Sour Water	556.9			0.8	0.5	558.2

Total Sour Water

Component	H ₂ O	H ₂ S	NH ₃	NH ₄ HS	NH ₄ OH	Total
Total Sour Water	15556			328	1015	16898

2.0 PROCESS DESCRIPTION

2.1 Hydrotreater (Area 100)

Operating conditions for the hydrotreater were provided to Lummus by Amoco and these conditions are presented in Table 2.1. The basic processing step selected was the expanded bed hydrotreater (LC Fining) system. Due to the extremely high exothermic heat of reaction it was necessary to use 3 reactors in with interstage cooling. Referring to drawing D5571-10101 and the material balance printouts (Appendix A) the flow is as follows:

- . Feed Tar Oil and Phenol are charged into the hydrotreater from day tanks FA-101 and FA-102 through charge pumps GA-101 and GA-102 and preheater exchanger EA-101.
- . The preheated charge oil is combined with feed hydrogen gas (at 576°F) from heater BA-101. Preheat of the oil is limited to 505°F to prevent cracking. The preheated mixture is then charged to the first reactor DC-101A.
- . The expanded bed reactor DC-101A approaches isothermal conditions in which the heat of reaction is used to heat the feed up to 700°F.
- . The effluent from DC101A is cooled in exchanger EA-101 and combined with recirculating hydrogen from recycle hydrogen gas compressor GB-102. The combined mixture (which has been cooled to 486°F) is charged into the second reactor where the heat of reaction increases the temperatures to 700°F.
- . The effluent from DC101B is cooled in exchanger EA-104 and combined with recirculating hydrogen gas from recirculating compressor GB-102. This mixture is charged into the third reactor DC-101C where its temperature rises from 531°F to 700°F.
- . The effluent from DC101C goes to the high temperature/high pressure separator FA-103. Hot liquid from FA-103 flows to the hydrotreater fractionation DA-101. The vapors from FA-103 flows through exchangers EA-102 and EA-105 and then through air cooler EC-101. Process water is injected prior to EC-101 to convert the H₂S and NH₃ in the gas to an aqueous NH₄OH/NH₄HS solution.
- . Exchangers EA-104 and EA-105 are part of a circulating hot oil belt which allows for the generation of steam from waste heat in the high pressure loop without having the problem of a hydrocarbon leak from the high pressure system into the steam system.

2.1 Hydrotreater - Cont'd

- . The cooled gas then passes into the High Pressure/Low Temperature Separator FA-104 where hydrogen rich gas is taken as an overhead product. A purge stream of this high pressure gas is taken (to purge H₂ and light gases from the loop) and sent to the Rectisol Unit 1400 in the SNG plant to recover the hydrogen in the purge gas. The remaining gas is recirculated to reactors DC-101B and DC-101C.
- . The water phase from separator FA-104 goes to the PHOSAM Unit in the SNG plant to recover the H₂S and NH₃.
- . The hydrocarbon phase from separator FA-104 is preheated in exchanger EA-105 and is combined with the hot liquid from FA-103 and charged to the HDT Fractionator DA-101. Fractionator DA-101 produces 500°F+ product (which is sent to hydrocracking, area 200), JP-4 (which is sent to storage), and unstabilized naphtha (which is sent to the naphtha stabilizer in the hydrocracking area 200).
- . Catalyst is replaced every third day in each reactor so that one reactor is receiving and withdrawing catalyst each day. Catalyst is added and replaced by the catalyst handling system.
- . Waste heat is converted to 100 psig saturated steam in exchangers EA-107 and EA-108. This steam is used for stripping in DA-101 and in the hydrocracking area 200 for stripping steam. There is an excess of about 5280 #/hr which is exported to the SNG steam system.

Table 2.1 Hydrotreater Conditions

Case 1 Maximum JP-4 Operation

Reactor Type	Expanded Bed
Number of Reactors	3
Reactor Temperature	700°F
Temp. rise/stage	225°F max.
Ratio of H ₂ in Feed to Chemical H ₂	2.0 min.
Catalyst Replacement	0.18 #/Bbl

2.2 Hydrocracker (Area 200)

Operating conditions for the hydrocracker were provided to Lummus by AMOCO and these conditions are presented in Table 2.2. The basic processing step selected was a 5 bed fixed bed reactor system with recycle of unconverted 500°F+ material. Beds 1 and 2 use a catalyst that is most active for sulfur, nitrogen and oxygen removal while beds 3,4,5 use a catalyst that is most active for hydrocracking. Referring to drawing D5571-10201 and the material balance printouts (Appendix B) the flow is as follows:

- Hydrotreated 500°F+ material from the hydrotreater (Area 100) enters the system from day tank FA-201 through feed pump GA-201 and is preheated in exchanger EA-201. The preheated feed is combined with unconverted bottoms from fractionator DA-201 (approximately 35% of the feed is recycled). The combined oils are then mixed with hot hydrogen coming from heater BA-201 and charged to the reactor.

The combined feed to the first bed in the reactor is 670°F.

- In the first bed the temperature rises to about 685°F. Quench hydrogen is added to cool the effluent from the first bed to about 652°F. In each of the remaining beds quench hydrogen is added to cool the beds. The inlet and outlet temperatures from each bed are as follows:

	Inlet	Outlet
Bed 1	675	685
Bed 2	652	683
Bed 3	653	682
Bed 4	653	688
Bed 5	656	696

- The reactor effluent is cooled in EA-201 and passes into the high temperature/high pressure separator FA-202. Vapors from FA-202 are cooled in EA-202 and then air condenser EC-201. Water is injected into the condenser EC-201 to dissolve H₂S and NH₃ into a NH₄OH/NH₄HS solution. This solution is sent to the PHOSAM unit in the SNG plant.
- The cooled vapors pass into separator FA-203 and the overhead hydrogen rich gas is divided with the major portion being used as recycle gas to the reactors via compressor GB-202 and heater BA-201. A small portion of the gas is purged from the system as high pressure purge gas which goes to the Rectisol unit in the SNG plant.

2.2 Hydrocracker (Area 200)

- . The hydrocarbon phase from separator FA-203 is heated in exchanger EA-202, combined with the hot oil from separator FA-202 and charged to fractionator DA-201.
- . Fractionator DA-201 produces unstabilized naphtha (which is charged to naphtha stabilizer DA-203), JP4 (which is sent to product storage after cooling), 500⁰F+ oil (which is recycled to reactor DC-201) and fuel gas (which flows to the boiler in the SNG plant).
- . Unstabilized naphtha from DA-201 is combined with unstabilized naphtha from area 100 and charged to naphtha stabilizer DA-203 after being preheated in exchanger EA-205. Heat for reboiling the naphtha stabilizer is obtained by heat exchange with the hot jet fuel product.
- . Fuel gas from the naphtha stabilizer is combined with fuel gas from FA-206 and is routed to the SNG boilers.
- . The stabilized naphtha is cooled and sent to storage. It is also sent to the SNG boilers to be used as fuel.

Table 2.2 Hydrocracker Conditions

Reactor Conditions	5 Fixed Beds
Catalyst, % of Total & Type	
Bed 1	10%, HDS/HDN/HC
Bed 2	22.5%, HDS/HDN/HC
Beds 3-5	22.5% HC
WHSV, hr ⁻¹	1.1
Average Reactor Temp.	670 ⁰ F
Temperature Increase	
Bed 1	25 ⁰ F
Bed 2-5	50 ⁰ F
Heat of Reaction	20,000 BTU/#Mole H ₂
Reactor Pressure	
Inlet	1200 psig
Outlet	1175 psig
Recycle Rate H ₂	15,000 scf/Bbl
Conversion/Pass	65%
Catalyst Replacement	3 years @ \$6/#

2.3 PSA Hydrogen Unit & Recompression (Area 300)

- 2.3.1 Hydrogen for both the hydrotreater and the hydrocracker will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig
Temp.	65 °F
Composition	mol%
H2	63.19
CO	18.61
CO2	1.48
CH4	16.21
C2H6	0.31
COS, H2S, CS2	< 0.01
N2 + Ar	0.19
H2O	< 0.01

The PSA unit selectively absorbs all components except H2 and produces a 99.99% vol. purity stream at about 345 psig and 80°F. The other components are available as a purge gas having the following properties.

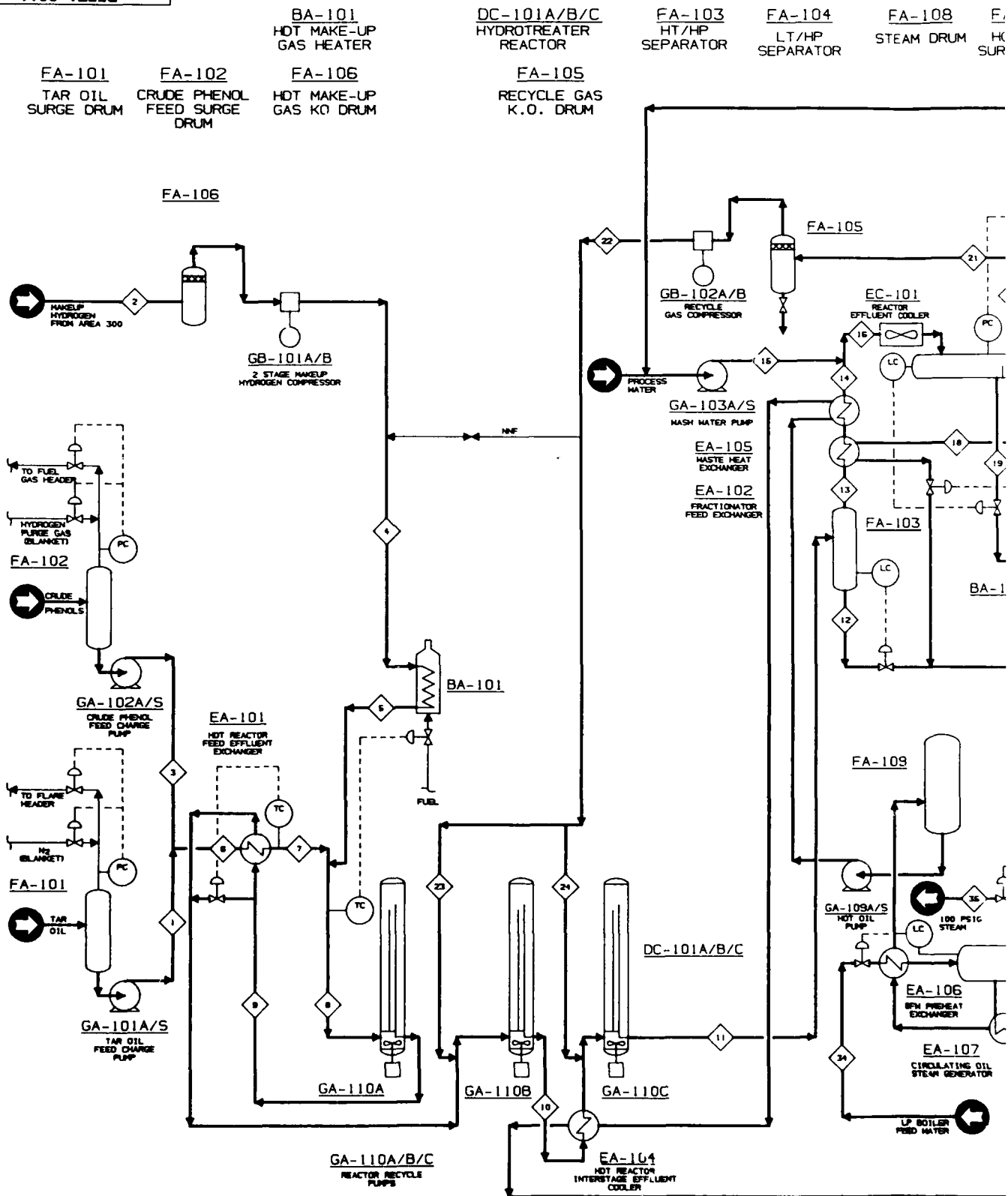
Pressure	5 psig
Temperature	100°F
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

At the conditions given a 10 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

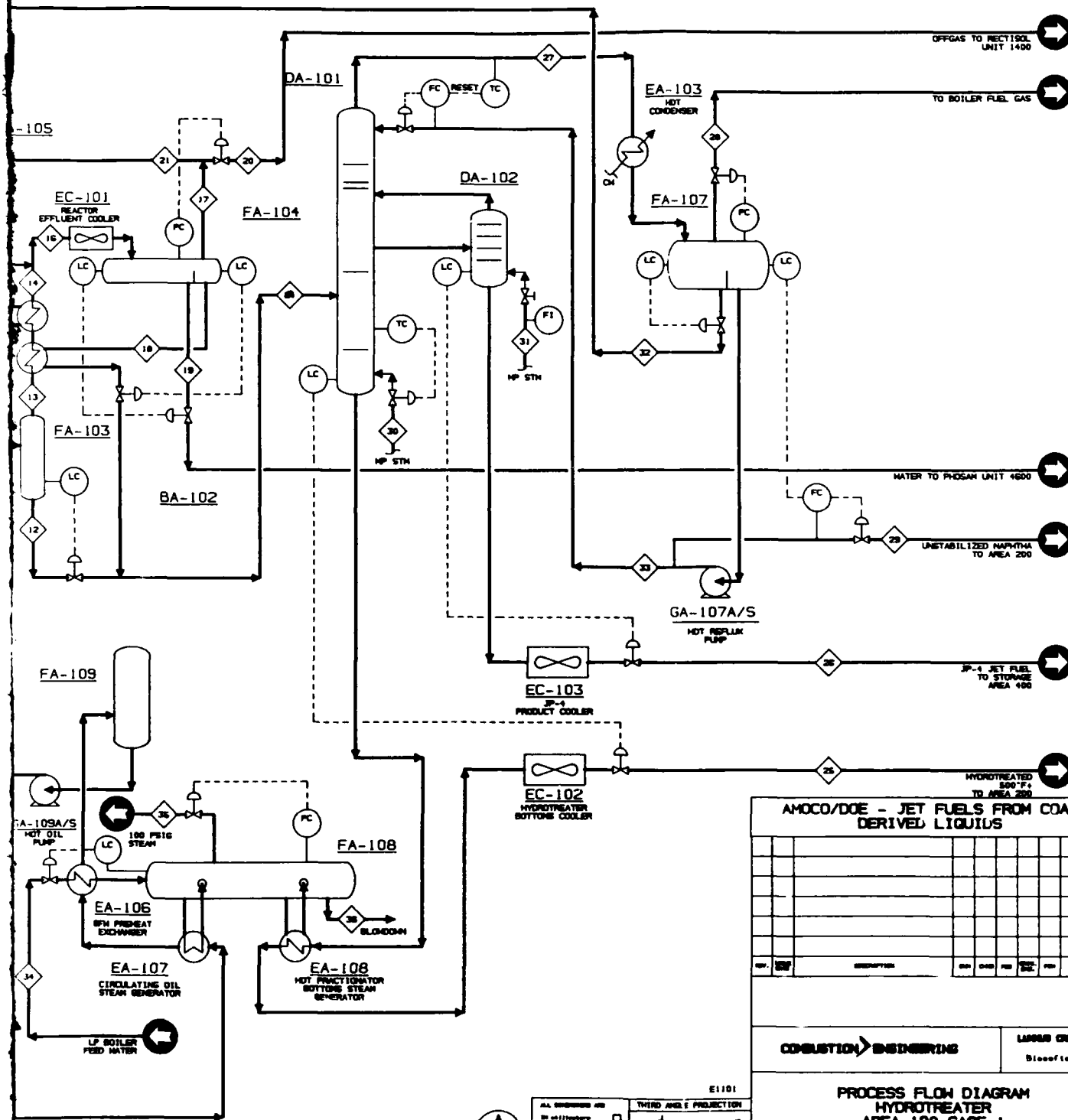
The system uses 10 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continuously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-10301 presents a schematic of a Union Carbide Polybed PSA unit.

- 2.3.2 The purge gas is recompressed to 375 psia and sent to the methanation unit of the SNG plant.

1100-17590



FA-107
FRACTIONATOR
REFLUX DRUM



AMOCO/DOE - JET FUELS FROM COAL
DERIVED LIQUIDS

[illegible]

CONSTRUCTION & ENGINEERING

LEONARD CRIST INC.
Blossfield, NJ

PROCESS FLOW DIAGRAM HYDROTREATER AREA 100 CASE 1

SCALE	DWG. NO.	D5571-10101

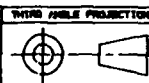


FA-207
STABILIZER
REFLUX DRUM

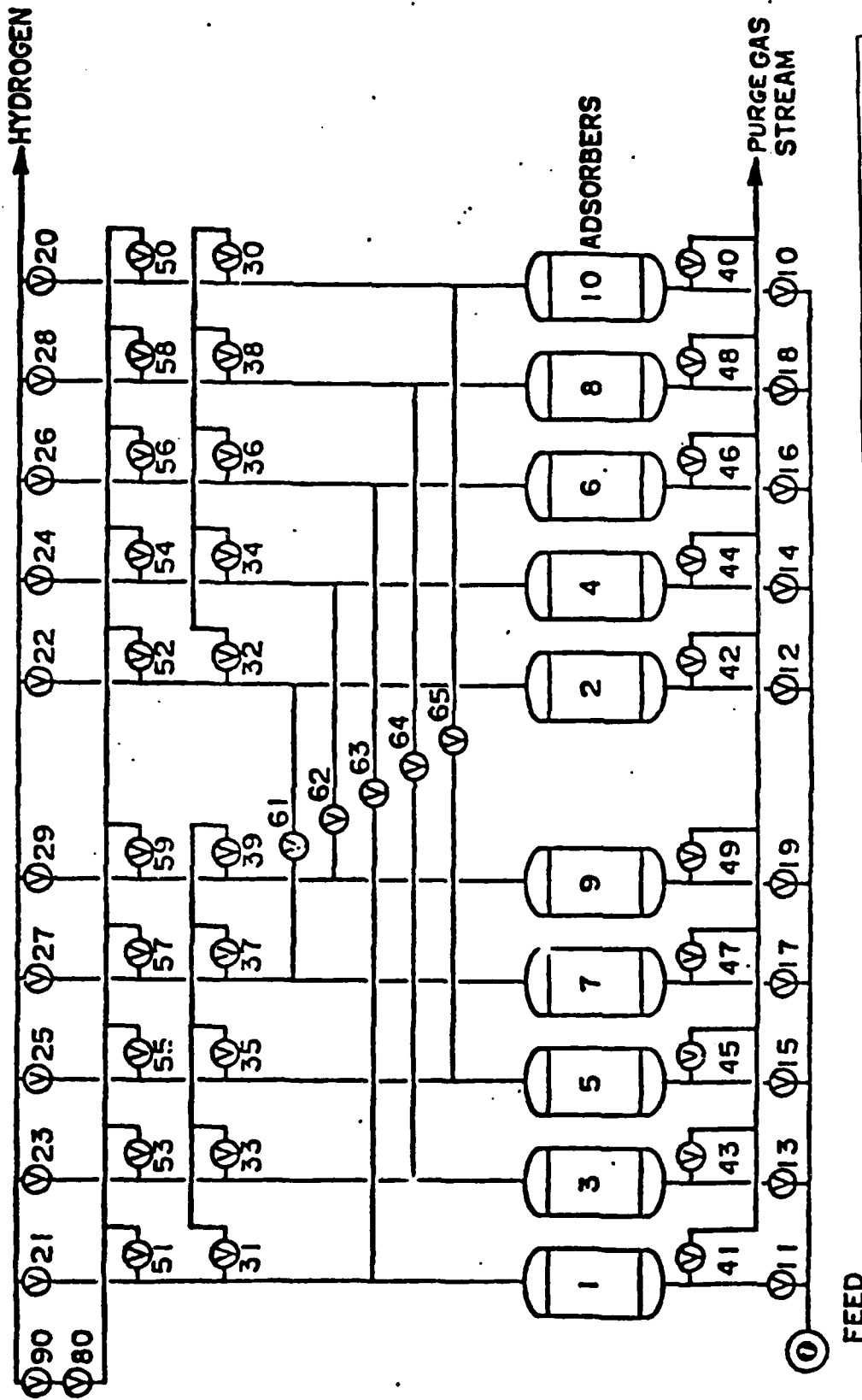
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LAGOS CRIST INC.
Bloomfield, NJ

SCALE	DWG. NO.	D5571-10201



E1004



TYPICAL ARRANGEMENT:

NUMBER OF ABSORBERS FOR THIS CASE = 10

[illegible]

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.1 Equipment List

CASE 1 - MAXIMUM JP-4

AREA 100 - HYDROTREATER

TAG NO. - DESCRIPTION

BA-101	HDT Makeup Gas Heater
DA-101	HDT Fractionator
DA-102	JP-4 Stripper
DC-101A,B,C	Hydrotreater Reactors
EA-101	HDT Reactor Feed/Effl. Exch.
EA-102	Fract. Feed Exch.
EA-103	HDT Condenser
EA-104	HDT Reactor Int. Stg. Clr.
EA-105	Waste Heat Exchanger
EA-106	BFW Preheat Exch.
EA-107	Circulating Oil Stm. Gen.
EA-108	HTD Fract. Btms. Stm. Gen.
EC-101	Reactor Effl. Cooler
EC-102	Hydrotreater Btms. Cooler
EC-103	JP-4 Product Cooler
FA-101	Tar Oil Surge Drum
FA-102	Crude Phenol Surge Drum
FA-103	HT/HP Separator
FA-104	LT/HP Separator
FA-105	Recycle Gas KO Drum
FA-106	HDT Makeup Gas Ko Drum
FA-107	Fractionator Reflux Drum
FA-108	Steam Drum
FA-109	Hot Oil Surge Drum
GA-101A/S	Tar Oil Feed Charge Pump
GA-102A/S	Crude Phenol Feed Charge Pump
GA-103A/S	Wash Water Pump
GA-107A/S	HDT Reflux Pump
GA-109A/S	Hot Oil Pump
GA-110A/B/C	Reactor Recycle Pump
GB-101A/B	H ₂ Makeup Compr.
GB-102A/B	Recycle Gas Compr.

CASE 1 - MAXIMUM JP-4 - Cont'd

AREA 200 - HYDROCRACKER

TAG. NO. DESCRIPTION

BA-201	HDC Reactor Gas Heater
DA-201	HDC Fractionator
DA-202	HDC JP-4 Stripper
DA-203	Naphtha Stabilizer
DC-201	HDC Reactor
EA-201	HDC Reactor Feed/Effl. Exch.
EA-202	HDC Fract. Feed Exch.
EA-203	HDC Fract. Condenser
EA-204	Stabilizer Reboiler
EA-205	Stabilizer Feed Exch.
EA-206	Stabilizer Condenser
EA-207	Naphtha Cooler
EC-201	Reactor Effl. Cooler
EC-202	JP-4 Product Cooler
FA-201	HDC Feed Surge Drum
FA-202	HT/HP Separator
FA-203	LT/HP Separator
FA-204	Recycle Gas KO Drum
FA-205	Makeup Gas KO Drum
FA-206	HDC Fract. Reflux Drum
FA-207	Stabilizer Reflux Drum
FA-208	Fuel Oil Day Tank
GA-201A/S	HDC Feed Pump
GA-202A/S	Wash Water Pump
GA-203A/S	HDC Recycle Pump
GA-204A/S	HDC Naphtha Pump
GA-207A/S	Stabilizer Btms Pump
GA-208A/S	Stabilizer Reflux Pump
GA-209A/S	Fuel Oil Pump
GB-201A/B	Makeup Gas Compr.
GB-202A/B	Recycle Gas Compr.

AREA 300 - PSA HYDROGEN UNIT & RECOMPRESSION

FA-301	Purge Gas Surge Drum
GB-301	Purge Gas Compressor
PA-301	PSA Hydrogen Unit Package

CASE 1 - MAXIMUM JP-4 - Cont'd

TAG NO. DESCRIPTION

AREA 400 - STORAGE AREA

FB-401 Jet Fuel Storage Tank
FB-402 Naphtha Storage Tank
FB-403 Fuel Oil Storage Tank

GA-401A/S Tar/Tar Oil Feed Pump
GA-402A/S Crude Phenol Feed Pump
GA-403A/S Fuel Oil Transfer Pump
GA-404A/S Naphtha Transfer Pump

AREA 500 - CATALYST HANDLING

TAG NO. DESCRIPTION

FA-501 Catalyst Oil Drum
FA-502 Catalyst Storage Hopper
FA-503 Catalyst Transfer Vessels
FA-504 Spent Catalyst Vessel

FL-501 Catalyst Screen

GA-501A/S Catalyst Transfer Pump
GA-502A/S Catalyst Oil Pump

3.2 Cost Estimate

3.2.1 Basis of Estimate

The estimate is an equipment factored type estimate using the equipment sizes & specifications developed for this project. The equipment unit pricing is based on return data for high pressure equipment purchased for various hydrotreater/hydrocracker projects. The unit pricing is some what conservative compared to world wide markets of 2-3 years ago, however, the exchange rate decline during this period will lead to higher purchase prices.

The commodity materials & subcontracts are ratioed from the equipment costs using factors considering the high pressure processing, the size of the units, and the location of the plant.

The labor and indirects also are factored considering process, sizing, and location.

Engineering costs are based on the equipment count times the historical number of manhours per equipment item, and the current average engineering selling rate.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2 Estimate Summary

(Thousands of \$)

Case 1

Area 100 Hydrotreater	\$23,836
Area 200 Hydrocracker	11,598
Area 300 PSA & Recompression	9,600
Area 400 OSBL	5,111
Area 500 Catalyst Handling	1,285
Total	<u>\$51,430</u>

3.2.3 Estimate Breakdown (Area 100) All Values in Thousands

	<u>Equipment</u>	<u>\$ Value</u>	<u>% Comm</u>	<u>\$ Comm.</u>
<u>Items</u>	<u>Type</u>			
1	Heaters	80	120	96
2	Towers	55	140	77
	Internals	7	-	-
3	Reactors	2025	65	1316
12	Exchangers	625	70	438
3	Air Coolers	99	100	99
9	Vessels	429	85	365
13	Pumps	881	80	705
4	Compressors	2050	60	1230
	Special	-		
47	Total	\$6251		\$4326
	Equipment		6251	
	Commodities		4326	
	Labor		3221 10% Equip. 60% Comm.	
	Indirects		3221 100%	
	Office		<u>2844</u> 47 pcs x 1100 x \$55-	
	Subtotal		<u>19,863</u>	
	Contingency		<u>3,973</u> 20%	
	Total		\$23,836	

3.2.3 Estimate Breakdown - Cont'd

Area 200

	<u>Equipment</u>	<u>\$ Valve</u>	<u>% Comm</u>	<u>\$ Comm.</u>
<u>Items</u>	<u>Type</u>			
1	Heaters	175	120	210
3	Towers	52	140	73
	Internals	8	-	-
1	Reactors	400	85	340
8	Exchangers	181	100	181
2	Air Coolers	57	110	63
8	Vessels	261	100	261
14	Pumps	162	120	194
4	Compressors	800	100	800
	Special	-		
41	Total	\$2096		\$2122
	Equipment		2096	
	Commodities		2122	
	Labor		1483	10% Equip. 60% Comm.
	Indirects		1483	100%
	Office		2481	41 pcs x 1100 x \$55-
	Subtotal		9665	
	Contingency		1933	20%
	Total		\$11,598	

Area 300

PSA-unit 20 mm SCFD budget quote \$3500

PSA unit 17 mm SCFD 3000

Installation 50% 1500

Subtotal \$4500

Compressor	3900	1950 x 2.0 T.I.C.
Drum	<u>300</u>	100 x 3.0 T.I.C.

Subtotal \$8700

Contingency 900 10%

Total \$9600

3.2.3 Estimate Breakdown - Cont'd

Area 400

Equipment & Value

Tankage		MTLS/C	805
Pumps		MTL	44
Yard Piping	28,400 LF	MTL	\$ 475
Labor	22,000 Hrs x \$50/hr	S/C	\$1,100
Pipe Insulation		S/C	160
Tracing		S/C	210
Excavation for U/G Pipe	55,000 Y ³	S/C	275
Rack	150 Ton Steel	S/C	300
Rack	FDN-500 Y ³	S/C	200

Equipment Related Commodities

Insulation	18000 Ft ² x \$10	S/C	180
FDNS, Instr, Elec, Paint - 20% x Equip.		MTL	170
S/C Labor @ 100% MTL		S/C	170
Total			\$4,089

Contingency 25% \$1,022

Total \$5,111

Area 500

	<u>Equipment</u>	<u>\$ Value</u>	<u>% Comm</u>	<u>\$ Comm.</u>
<u>Items</u>	<u>Type</u>			
4	Vessels	105	120	126
4	Pumps	48	120	58
8	Total	\$153		\$184

Equipment	153
Commodities	184
Labor	125 10% Equip. 60% Comm.
Indirects	125 100%
Office	484 8 pcs x 1100 x \$55-
Subtotal	\$1071
Contingency	214 20%
	\$1285

4.0 OPERATING COSTS

4.1 Operating Labor

It is estimated that it will require 7 men/shift to operate the plant broken down as follows:

Foreman	1	
Control Room	1	
HDT Operator	2	
HCR Operator	2	
PSA & relief man	1	
	<u>7</u>	Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 5 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	7 positions x 4 people/position -	28
Supervisor & Admin.		5
QC Technician		1
Maintenance		5
Other (Stores or Janitorial)		1
Total		<u>40</u>

4.2 Utilities

The following utilities have been estimated from the computer simulations:

<u>Utility</u>	<u>Consumption</u>	<u>Cost</u>	<u>\$/SD</u>
#6 Fuel Oil	3372 BPSD	\$16/Bbl (a)	53952
SNG equivalent	4.81 MMSCFD	\$3.80/MM Btu (b)	17912
of Syn Gas & Purge Gas			
Cooling Water	2400 GPM	\$0.155/MGal (c)	536
Power	7100 kW	\$0.04/kWh (c)	6816
Process Water	18.5 GPM	\$0.45/MGal (c)	12

(a) Cut of 1% sulfur \$6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

(b) Memo from D. Daley of Burns & Roe to L. Lorenzo of DOE dated Oct. 20, 1987, reference DPD-87-863.

(c) ANG utility cost information dated 5/87.

4.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

<u>Catalyst</u>	<u>Use</u>	<u>Cost</u>	<u>\$/SD</u>
HDT Cat.	0.18 #/Bbl	\$3.00/#	2218
HCR Cat.	0.013 #/Bbl	\$6.00/#	96
Inhibitors	50 PPM	\$10/Gal	86
			<u>2400</u>

4.4 Maintenance Supplies

Maintenance supplies for hydrotreating operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units. On this basis the maintenance supplies would be

$$0.00005 \times 51,430,000 = \$2571/\text{SD}$$

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the production of JP-4 are proposed to be located to the east of the Rectisol Unit and Main Control Room of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 300' x 220' will be surrounded by an access road and will be divided by a central east-west road. Areas 100 & 500 will be located to the north and Areas 200 & 300 to the south.

A diked storage tank area approx. 360' x 265' will be required for product and fuel oil storage and is proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

5.2 Unit Tie-Ins

Approximately 2000 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

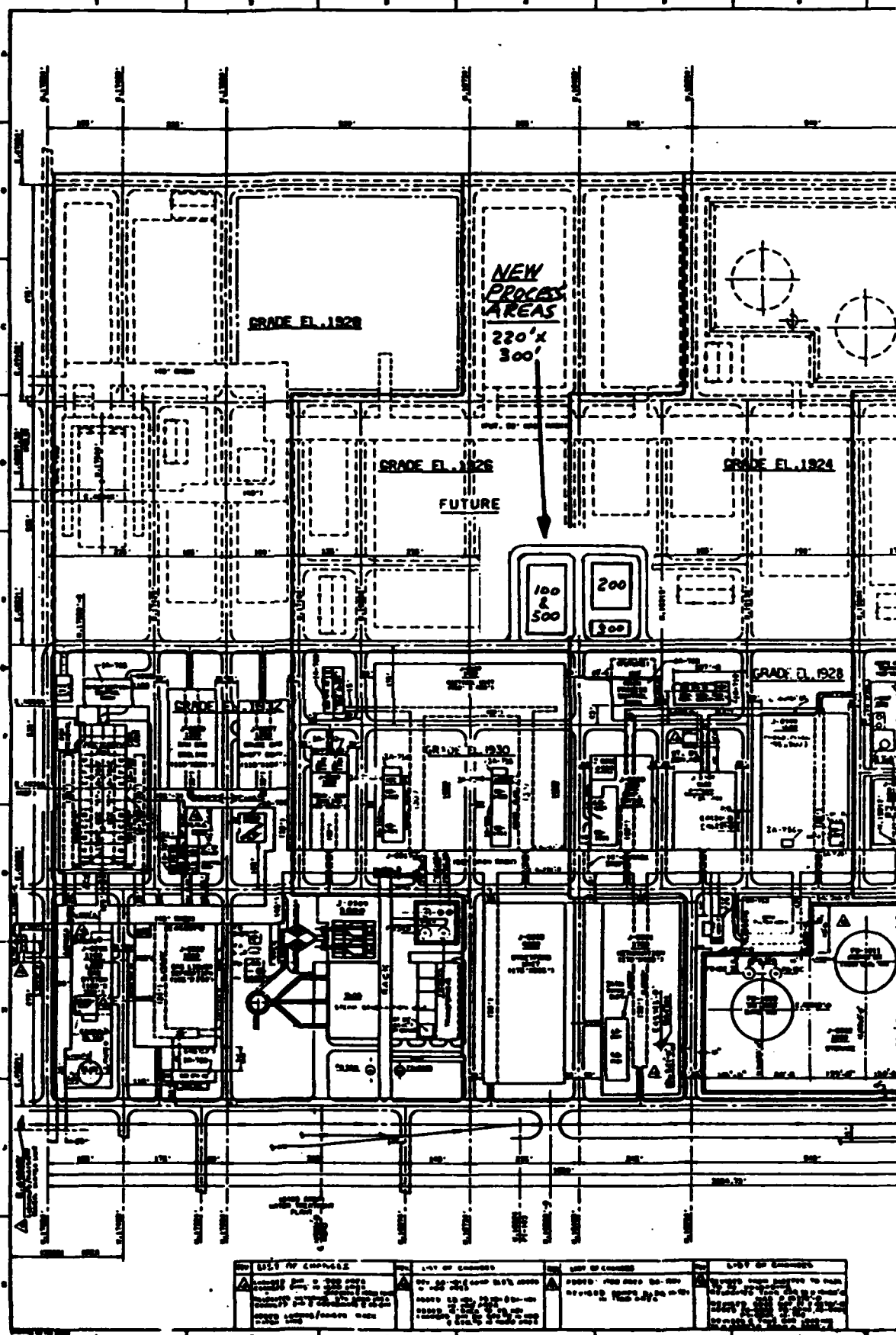
New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

A summary of the lines is shown in table 5.1.

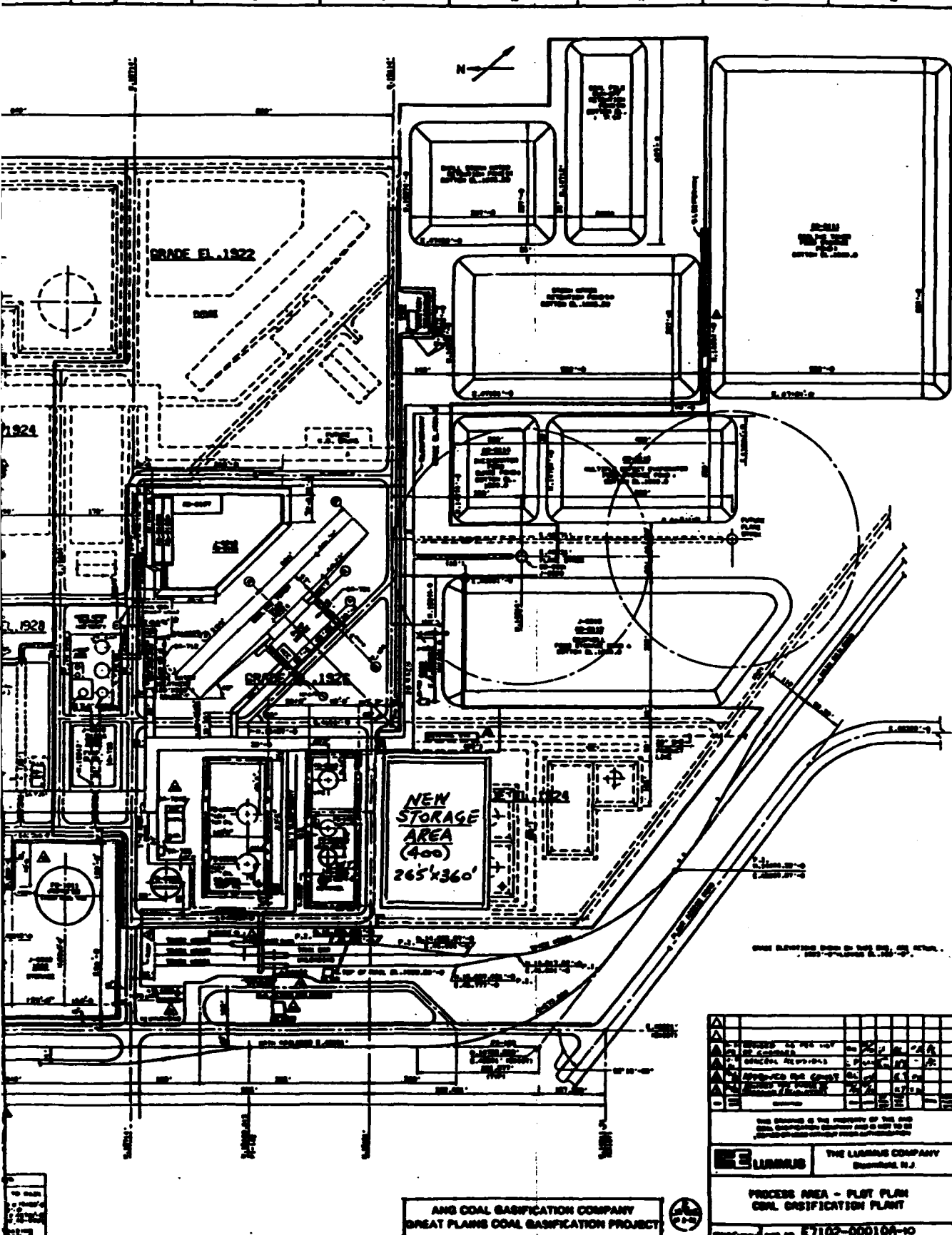
TABLE 5.1
INTERCONNECTING PIPING

<u>EST. SIZE</u>	<u>SERVICE</u>	<u>TO/FROM</u>
4"	Tar/Tar Oil (Elec. Tr.)	Storage
2"	Crude Phenol (Elec. Tr.)	Storage
4"	JP-4 Product	Storage
1 1/2"	Naphtha Product	Storage
18"	Wet Flare (Trace)	Flare
8"	Synthesis Gas	PSA/Rectisol
6"	Purge Gas	Methanation/PSA
2"	Off Gas	Rectisol/HDT, HDC
2"	Nitrogen	Main Rack
2"	Plant Air	"
2"	Instr. Air	"
2"	Raw Water (Elec. Tr.)	"
6"	M.P. Steam	"
1 1/2"	Stm Cond.	"
1 1/2"	BFW	"
1 1/2"	Boiler B.D.	"
12"	C. W. Supply & Return	"
2"	Waste Water	Phosam/HDT, HDC
6"	Fuel Oil	Exist TKS/New TKS.
15"	Storm Sewer (9' deep)	Storm Basin
15"	Oily Water Sewer (9' deep)	8100/Process Unit
6"	Sanitary Sewer (9' deep)	8400/Process Unit
10"	Fire Water	Ring Headers





SECTION A	SECTION B	SECTION C	SECTION D
1. 100' x 500' AREA 2. 200' x 300' AREA 3. 300' x 200' AREA 4. 400' x 100' AREA 5. 500' x 100' AREA	1. 100' x 500' AREA 2. 200' x 300' AREA 3. 300' x 200' AREA 4. 400' x 100' AREA 5. 500' x 100' AREA	1. 100' x 500' AREA 2. 200' x 300' AREA 3. 300' x 200' AREA 4. 400' x 100' AREA 5. 500' x 100' AREA	1. 100' x 500' AREA 2. 200' x 300' AREA 3. 300' x 200' AREA 4. 400' x 100' AREA 5. 500' x 100' AREA



CASE 1

VERSION 2.01
SIMULATION SCIENCES INC.
PROJECT GP JET FUELS
PROBLEM C10100

PROCESS
SOLUTION

PAGE A-1
CASE I
AREA 100

PAGE 19
MAX JPL
OCT 16 1987

HYDROTREATER

STREAM NO.	1	2	3	4	5	6
STREAM NAME	FED TAP LIQUID	MAKEUP VAPOR	CP PHENOLS LIQUID	NET MAKEUP VAPOR	H2 FR B101 VAPOR	COMBINE FEED LIQUID
TEMPERATURE, DEG F	154.0000	70.0000	120.0000	393.0950	576.2288	147.7182
RATE LB MOLS/MR	504.247	1400.0000	141.7249	1799.9998	1799.9998	446.0147
RATE LB /MR	47619.9663	3631.3228	14212.9263	3631.3228	3631.3228	61832.9141
ENTHALPY HP BTU /MR	2.0423	-3.7421	0.4788	0.1852	2.5813	2.5211
ENTHALPY BTU /LB	42.8877	-1030.5193	33.6883	51.0032	683.2996	40.7739
MOLECULAR WEIGHT....	156.4980	2.0174	100.2522	2.0174	2.0174	138.6342
*** VAPOR PHASE ***						
RATE LB /MR	0.0000	3631.3228	0.0000	3631.3228	3631.3228	0.0000
STD. RATE MP FT3/DAY	0.00	16.39	0.00	16.39	16.39	0.00
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4475	3.4485	0.0000
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	2.0174	0.0000

*** LIQUID PHASE ***

RATE LB /MR	47619.9668	0.0000	14212.9863	0.0000	0.0000	61832.9141
ACT. RATE BBL/DAY	3261.67	0.00	929.35	0.00	0.00	4195.35
STD. LV RATE BBL/MR	132.25	0.00	34.20	0.00	0.00	171.06
CP, BTU /LB F	2.4043	0.0000	0.4025	0.0000	0.0000	0.4044
MOLECULAR WEIGHT....	156.4980	0.0000	100.2522	0.0000	0.0000	138.6342
ACT. DENS LB /FT3	62.4044	0.0000	65.3727	0.0000	0.0000	63.0005
STD. API GRAVITY....	6.5894	0.0000	1.5224	0.0000	0.0000	5.4247

*** DRY BASIS ***

RATE LB /MR	46667.9760	0.0000	13573.0332	0.0000	0.0000	60240.9688
MOLECULAR WEIGHT....	185.6022	0.0000	127.7966	0.0000	0.0000	168.4372
UOP K	9.9520	0.0000	9.9460	0.0000	0.0000	9.7261
FLASH POINT, DEG F	142.3670	0.0000	92.6111	0.0000	0.0000	119.6567
CRIT. TEMP, F	966.3024	0.0000	786.3197	0.0000	0.0000	912.8544
CRIT. PRES, PSIA	424.5659	0.0000	546.5264	0.0000	0.0000	461.0103

*** VAPOR PHASE ***

RATE LB /MR	0.0000	3631.3228	0.0000	3631.3228	3631.3228	0.0000
STD. RATE MP FT3/MR	0.00	0.63	0.00	0.68	0.68	0.00
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4475	3.4485	0.0000
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	2.0174	0.0000

VISCOSITY, CP

	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

*** LIQUID PHASE ***

RATE LB /MR	46667.9760	0.0000	13573.0332	0.0000	0.0000	60240.9688
ACT. RATE BBL/DAY	3195.29	0.00	845.02	0.00	0.00	4084.23
CP, BTU /LB F	0.3723	0.0000	0.3749	0.0000	0.0000	0.3889
MOLECULAR WEIGHT....	185.6022	0.0000	127.7966	0.0000	0.0000	168.4372
ACT. DENS LB /FT3	62.4312	0.0000	65.5543	0.0000	0.0000	63.0484
STD. API GRAVITY....	6.5112	0.0000	1.1127	0.0000	0.0000	5.3021

SIMULATION SCIENCES INC.
PROJECT GP JET FUELS
PROBLEM C1U100

HYDROTREATER

STREAM ID.	7	8	9	10	11	12
STREAM NAME	MX FD EFF	101A INLET	101A OUTLET	101B OUTLET	101C OUTLET	HOT SEP LIQ
STREAM PHASE	LIQUID	MIXED	MIXED	MIXED	MIXED	LIQUID
TEMPERATURE, DEG F	503.5925	472.5212	700.0000	700.0000	700.0000	699.9121
RATE LB MOLS/HR	446.0145	2249.0112	1724.6343	1686.4119	2182.7847	85.1379
RATE LB /HR	61832.9141	65606.2422	65463.9531	66547.2969	69449.2969	10872.4727
ENTHALPY MM BTU /HR	13.1170	15.5983	30.0528	32.8555	38.0253	3.8500
ENTHALPY BTU /LB	212.1343	233.2718	459.0744	493.7171	547.5342	354.1037
MOLECULAR WEIGHT	138.6343	29.1469	36.6000	39.5077	31.8164	127.7042
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	3441.2051	29373.4328	38375.8750	58528.3828	0.0000
STD. RATE MM FT3/DAY	0.00	16.38	13.86	13.30	19.10	0.00
CP, BTU /LB F	0.0000	1.6970	0.7879	0.7396	0.7558	0.0000
MOLECULAR WEIGHT	0.0000	4.5550	19.2990	26.2875	27.9085	0.0000

B-32

*** LIQUID PHASE ***						
RATE LB /HR	61832.9141	57023.0391	36090.0703	28171.4063	10919.9219	10872.4727
ACT. RATE BBL/DAY	4361.16	4040.74	3652.77	2990.44	1189.07	1184.28
STD. LV RATE BBL/HR	171.06	161.36	107.75	92.32	35.61	35.44
CP, BTU /LB F	0.5751	0.5505	0.6694	0.7192	0.7582	0.7379
MOLECULAR WEIGHT	138.6343	145.1591	135.4171	125.4331	127.5177	127.7642
ACT. DENS LB /FT3	54.3710	52.3237	42.2335	40.2685	39.2558	39.2434
STD. API GRAVITY	5.4247	5.5542	16.2450	23.6696	29.8915	29.8532
*** DRY BASIS ***						
RATE LB /HR	60240.9632	56767.1094	35702.9438	27655.0773	10731.0762	10685.0938
MOLECULAR WEIGHT	168.4372	149.9209	145.7165	141.1739	142.9699	142.9699
UOP K	9.7261	9.8267	10.2764	10.6338	10.9789	10.9781
FLASH POINT, DEG F	119.6567	25.3493	-19.0451	-50.2837	-54.2565	-54.1564
CRIT. TEMP, F	412.2564	751.4496	644.9341	605.8790	565.6893	570.5956
CRIT. PRES, PSIA	461.0103	419.8000	373.5549	374.0819	356.0148	355.9686
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	7105.1943	26169.4531	33370.1328	51425.2969	0.0000
STD. RATE MM FT3/HR	0.00	0.24	0.51	0.45	0.45	0.00
CP, BTU /LB F	0.0000	1.9196	0.2222	0.7718	0.7885	0.0000
MOLECULAR WEIGHT	0.0000	3.9930	19.4677	28.2321	30.1993	0.0000

VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	60240.9632	56767.1094	35702.9438	27655.0773	10731.0762	10685.0938
ACT. RATE BBL/DAY	423.20	4014.02	3591.31	2900.48	1159.10	1154.58
CP, BTU /LB F	0.5591	0.5476	0.6395	0.6669	0.7096	0.7096
MOLECULAR WEIGHT	168.4372	149.9209	145.7165	141.1739	142.9699	142.9699
ACT. DENS LB /FT3	54.3540	52.3338	42.4755	40.0444	36.5746	36.5590
STD. API GRAVITY	5.4247	5.5542	16.2450	23.6696	29.8915	29.8532

HYDROTREATER

STREAM ID.	15	16	17	18
STREAM NAME	H2O WASH EFF + M20	COLD SEP VAP	COLD SEP VAP	COLD SEP LIQ
STREAM PHASE	LIQUID	MIXED	VAPOR	LIQUID
TEMPERATURE, DEG F	699.9121	437.5931	120.0000	120.0000
RATE LB FCS/LHR	2097.6475	2097.6405	1260.6277	422.5313
RATE LB /HR	58575.3281	58575.4241	4063.7197	47032.2188
ENTHALPY MM BTU /HR	34.1750	13.0475	-1.8977	0.5580
ENTHALPY BTU /LB	583.4320	325.1774	-466.9748	11.8645
MOLECULAR WEIGHT....	27.9245	27.9245	3.2236	111.3106
*** VAPOR PHASE ***				
RATE LB /HR	58575.8281	20842.4219	4063.7197	0.0000
STD. RATE MM FT3/DAY	19.10	15.64	11.48	0.00
CP, BTU /LB F	0.7557	0.9528	2.2613	0.0000
MOLECULAR WEIGHT....	27.9245	12.1399	3.2236	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	0.0000	37733.3906	0.0000	47032.2188
ACT. RATE BBL/DAY	0.00	3965.44	0.00	4049.25
STD. LV RATE BBL/HR	0.00	130.05	0.00	165.76
CP, BTU /LB F	0.0000	0.6447	0.0000	0.4640
MOLECULAR WEIGHT....	0.0000	99.0930	0.0000	111.3106
ACT. DENS LB /FT3	0.0000	40.6708	0.0000	49.6492
STD. API GRAVITY....	0.0000	39.0882	0.0000	42.9141
*** DRY BASIS ***				
RATE LB /HR	0.0000	36473.8750	0.0000	47017.0154
MOLECULAR WEIGHT....	0.0000	117.3274	0.0000	111.4973
UOP K	0.0000	11.1938	0.0000	11.2645
FLASH POINT, DEG F	0.0000	-61.5335	0.0000	-77.8154
CRIT. TEMP, F	0.0000	525.6302	0.0000	549.8193
CRIT. PRES, PSIA	0.0000	407.4491	0.0000	478.1943
*** VAPOR PHASE ***				
RATE LB /HR	51471.2513	14997.3945	4044.3599	0.0000
STD. RATE MM FT3/DAY	0.45	0.53	0.48	0.00
CP, BTU /LB F	0.7883	1.0320	2.2688	0.0000
MOLECULAR WEIGHT....	30.2189	10.7709	3.2109	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	0.0000	36473.8750	0.0000	47017.0154
ACT. RATE BBL/DAY	0.00	3562.36	0.00	4048.22
CP, BTU /LB F	0.0000	0.6263	0.0000	0.4638
MOLECULAR WEIGHT....	0.0000	117.3274	0.0000	111.2645
ACT. DENS LB /FT3	0.0000	43.3665	0.0000	49.6461
STD. API GRAVITY....	0.0000	40.9905	0.0000	42.9551

HYDROTREATER

STREAM ID.	10	20	21	22	23	24
STREAM NAME	SOUR M20 PURGE GAS LIQUID	RECYCLE GAS VAPOR	REC COMP DIS VAPOR	101B REC GAS VAPOR	101C REC GAS VAPOR	
TEMPERATURE, DEG F	120.0000	120.0000	120.0000	120.0000	120.0000	120.0000
RATE LB POLS/MR	24.4908	1236.1367	1236.1384	336.1385	899.9999	
RATE LB /MR	78.9478	3994.7710	3994.7793	1083.5659	2901.2119	
ENTHALPY MM BTU /MR	-0.0369	-1.8608	-1.7003	-0.4624	-1.2380	
ENTHALPY BTU /LB	-466.9752	-466.9749	-426.7032	-426.7032	-426.7032	
MOLECULAR WEIGHT....	3.2236	3.2236	3.2236	3.2236	3.2236	
*** VAPOR PHASE ***						
RATE LB /MR	78.9478	3994.7710	3994.7793	1083.5659	2901.2119	
STD-RATE MM FT3/DAY	U-22	11.26	11.26	3.06	8.20	
CP, BTU /LB F	2.2603	2.2603	2.2582	2.2582	2.2582	
MOLECULAR WEIGHT....	3.2236	3.2236	3.2236	3.2236	3.2236	

*** LIQUID PHASE ***						
RATE LB /MR	0.0000	0.0000	0.0000	0.0000	0.0000	
ACT-RATE BBL/DAY	0.00	0.00	0.00	0.00	0.00	
STD. LV RATE BBL/MR	0.00	0.00	0.00	0.00	0.00	
CP, BTU /LB F	0.0000	0.0000	0.0000	0.0000	0.0000	
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	0.0000	0.0000	
ACT-DENS LB /FT3	0.0000	0.0000	0.0000	0.0000	0.0000	
STD. API GRAVITY....	0.0000	0.0000	0.0000	0.0000	0.0000	

*** DRY BASIS ***						
RATE LB /MR	0.0000	0.0000	0.0000	0.0000	0.0000	
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	0.0000	0.0000	
UOP K	0.0000	0.0000	0.0000	0.0000	0.0000	
FLASH POINT, DEG F	U-0000	0.0000	0.0000	0.0000	0.0000	
CRIT. TEMP, PSIA	0.0000	0.0000	0.0000	0.0000	0.0000	
CRIT. PRES, PSIA	0.0000	0.0000	0.0000	0.0000	0.0000	
*** VAPOR PHASE ***						
RATE LB /MR	78.9478	3994.7710	3994.7793	1083.5659	2901.2119	
STD-RATE MM FT3/MR	U-01	0.47	0.47	0.13	0.34	
CP, BTU /LB F	2.2603	2.2603	2.2673	2.2673	2.2673	
MOLECULAR WEIGHT....	3.2236	3.2236	3.2236	3.2236	3.2236	

VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	

*** LIQUID PHASE ***						
RATE LB /MR	0.0000	0.0000	0.0000	0.0000	0.0000	
ACT-RATE BBL/DAY	U-00	0.00	0.00	0.00	0.00	
CP, BTU /LB F	0.0000	0.0000	0.0000	0.0000	0.0000	
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	0.0000	0.0000	
ACT-DENS LB /FT3	0.0000	0.0000	0.0000	0.0000	0.0000	
STD. API GRAVITY....	0.0000	0.0000	0.0000	0.0000	0.0000	

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	25	26	27	28	29
STREAM NAME	MDT JP-4	FRAC OVHD	MDT HC VAP	MDT NAPHTHA	
STREAM PHASE	LIQUID	VAPOR	VAPOR	LIQUID	
TEMPERATURE, DEG F	549.6777	329.8319	236.9774	90.0000	90.0000
PRESSURE, PSIA	64.0000	45.5000	90.0000	80.0000	80.0000
RATE, LB PLS/H	507.5550	294.0670	1010.1720	91.2511	44.8024
RATE, LB /HR	57904.6406	35947.9922	59035.8672	1727.7163	3209.6353
ENTHALPY MM BTU /HR	15.5373	4.0676	15.8779	0.3145	0.0086
ENTHALPY BTU /LB	337.4049	113.1529	268.9532	182.0228	2.6839
MOLECULAR WEIGHT	114.0598	122.2442	58.4414	18.9337	71.6398
*** VAPOR PHASE ***					
RATE, LB /HR	45181.0078	0.0000	59035.8672	1727.7163	0.0000
ACT. RATE FT3/SEC	20.87	0.00	21.61	1.84	0.00
STD. RATE MM FT3/DAY	4.25	0.00	9.20	0.83	0.00
CP, BTU /LB F	0.5800	0.0000	0.4839	0.5648	0.0000
MOLECULAR WEIGHT	105.5102	0.0000	58.4414	18.9337	0.0000
ACT. DENS LB /FT3	0.6547	0.0000	0.7590	0.2606	0.0000
COMPRESSIBILITY (Z)	0.9522	0.0000	0.9270	0.9854	0.0000
*** LIQUID PHASE ***					
RATE, LB /HR	8723.6426	35947.9922	0.0000	0.0000	3209.6353
ACT. RATE BBL/DAY	816.86	3534.36	0.00	0.00	318.73
STD. LV RATE BBL/H	26.62	124.85	0.00	0.00	12.98
CP, BTU /LB F	0.6210	0.5779	0.0000	0.0000	0.5048
MOLECULAR WEIGHT	205.9675	122.2442	0.0000	0.0000	71.6398
ACT. DENS LB /FT3	45.6503	43.4521	0.0000	0.0000	43.0434
STD. API GRAVITY	19.5504	40.4044	0.0000	0.0000	68.6129
*** DRY BASIS ***					
RATE, LB /HR	5723.1992	35772.2266	0.0000	0.0000	3208.8184
MOLECULAR WEIGHT	210.1013	125.8212	0.0000	0.0000	71.6941
UOP K	10.7600	11.2069	0.0000	0.0000	11.7920
FLASH POINT, DEG F	110.4238	48.4767	0.0000	0.0000	-74.4763
CRIT. TEMP, PSIA	904.1510	658.1058	0.0000	0.0000	422.6767
CRIT. PRES, PSIA	337.0012	436.6072	0.0000	0.0000	575.3866
*** VAPOR PHASE ***					
RATE, LB /HR	42978.8750	0.0000	56144.7109	1713.4653	0.0000
ACT. RATE FT3/SEC	20.34	0.00	17.95	1.83	0.00
STD. RATE MM FT3/DAY	0.17	0.00	0.32	0.03	0.00
CP, BTU /LB F	0.5807	0.0000	0.4878	0.5658	0.0000
MOLECULAR WEIGHT	107.6624	0.0000	66.0780	18.9417	0.0000
ACT. DENS LB /FT3	0.6689	0.0000	0.8688	0.2607	0.0000
COMPRESSIBILITY (Z)	0.9511	0.0000	0.9154	0.9853	0.0000
VISCOSITY, CP	0.0127	0.0000	0.0098	0.0110	0.0000
*** LIQUID PHASE ***					
RATE, LB /HR	5723.1992	35772.2266	0.0000	0.0000	3208.8184
ACT. RATE BBL/DAY	816.86	3523.01	0.00	0.00	318.67
CP, BTU /LB F	0.6210	0.5760	0.0000	0.0000	0.5047
MOLECULAR WEIGHT	210.1013	125.8212	0.0000	0.0000	71.6941
ACT. DENS LB /FT3	45.6503	43.4015	0.0000	0.0000	43.0431
STD. API GRAVITY	19.5511	40.4044	0.0000	0.0000	68.6278

VERSION 1
SIMULATION SCIENCES INC.
PROJECT GP JET FUELS
PROBLEM C10100

TM

PROCESS

SOLUTION

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	30	31	32	33
STREAM NAME	FRAC STR STM JP-4 STR STM	REFLUX	SCUR M20	LIQUID
STREAM PHASE	VAPOR	LIQUID	LIQUID	LIQUID
TEMPERATURE, DEG F	330.6417	331.3275	50.0000	90.0000
PRESSURE, PSIA	75.8998	95.5000	20.0000	80.0000
RATE, LB PCL/HR	72.1621	35.9149	158.9312	715.1871
RATE, LB /HR	1299.9590	1599.9992	2863.1460	51235.8438
ENTHALPY MM BTU /HR	1.3473	1.9043	0.1661	0.1375
ENTHALPY BTU /LB	1190.2109	1190.2109	52.0156	2.6339
MOLECULAR WEIGHT	18.0150	18.0150	18.0150	71.6398
*** VAPOR PHASE ***				
RATE, LB /HR	1299.9990	1599.9992	0.0000	0.0000
ACT-RATE, FT3/SEC	1.72	2.06	0.00	0.00
STD-RATE MP FT3/DAY	0.66	0.51	0.00	0.00
CP, BTU /LB F	0.5787	0.5805	0.0000	0.0000
MOLECULAR WEIGHT	18.0150	18.0150	0.0000	0.0000
ACT-DENS, LB /FT3	0.2100	0.2136	0.0000	0.0000
COMPRESSIBILITY (Z)	0.9453	0.9490	0.0000	0.0000
*** LIQUID PHASE ***				
RATE, LB /HR	0.0000	0.0000	2563.1460	51235.8438
ACT-RATE, BBL/DAY	0.00	0.00	197.08	5087.91
STD. LV RATE BBL/HR	0.00	0.00	3.19	207.15
CP, BTU /LB F	0.0000	0.0000	0.9978	0.5047
MOLECULAR WEIGHT	0.0000	0.0000	18.0150	71.6398
ACT-DENS, LB /FT3	0.0000	0.0000	42.1002	43.0454
STD. API GRAVITY	0.0000	0.0000	10.0635	68.6129
*** DRY BASIS ***				
RATE, LB /HR	0.0000	0.0000	0.0000	51222.9047
MOLECULAR WEIGHT	0.0000	0.0000	0.0000	71.6941
UOP K	0.0000	0.0000	0.0000	11.7920
FLASH POINT, DEG F	0.0000	0.0000	0.0000	-74.4764
CRIT. TEMP, F	0.0000	0.0000	0.0000	422.6769
CRIT. PRES, PSIA	0.0000	0.0000	0.0000	575.3867
*** VAPOR PHASE ***				
RATE, LB /HR	0.0000	0.0000	0.0000	0.0000
ACT-RATE, FT3/SEC	0.00	0.00	0.00	0.00
STD-RATE MM FT3/HR	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT	0.0000	0.0000	0.0000	0.0000
ACT-DENS, LB /FT3	0.0000	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z)	0.0000	0.0000	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***				
RATE, LB /HR	0.0000	0.0000	0.0000	51222.8047
ACT-RATE, BBL/DAY	0.00	0.00	0.00	5087.02
CP, BTU /LB F	0.0000	0.0000	0.0000	0.5046
MOLECULAR WEIGHT	0.0000	0.0000	0.0000	71.6941
ACT-DENS, LB /FT3	0.0000	0.0000	0.0000	43.0454
STD. API GRAVITY	0.0000	0.0000	0.0000	68.6129

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	1	2	3	4	5	6
STREAM NAME	HOT SSC+ LIQUID	MAKEUP H2 VAPOR	HDC RECYCLE LIQUID	NET MAKE-UP VAPOR	RX INLET MIXED	RX EFFLUENT VAPOR
TEMPERATURE, DEG F	200.0000	70.0000	310.0000	443.2653	67C.0000	696.1902
PRESSURE, PSIA	1240.0000	340.0000	1240.0000	1265.0000	120C.0000	1475.0000
RATE, LB PCL/HR	77.4363	250.0000	37.4100	250.0000	1461.0029	2367.8394
RATE, LB /HR	17056.6367	504.3504	8901.9961	504.3504	30291.7891	33940.1719
ENTHALPY MM BTU /HR	0.4604	-0.5197	1.8367	0.1261	12.6494	18.7233
ENTHALPY BTU /LB	26.9926	-1030.5193	206.3200	249.9401	417.5854	551.6570
MOLECULAR WEIGHT	220.3512	2.0174	237.6579	2.0174	2C.7336	14.3336
*** VAPOR PHASE ***						
RATE, LB /HR	C.0000	504.3504	0.0000	504.3504	15859.0449	33940.1719
ACT. RATE, FT3/SEC	0.00	1.18	0.00	0.55	4.03	7.17
STD. RATE MM FT3/DAY	0.00	2.29	0.00	2.28	12.65	21.57
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4758	1.0812	1.0240
MOLECULAR WEIGHT	0.0000	2.0174	0.0000	2.0174	11.4196	14.3336
ACT. DENS, LB /FT3	C.0000	0.1189	0.0000	0.2543	1.0935	1.3156
COMPRESSIBILITY (Z)	0.0000	1.0145	0.0000	1.0357	1.0338	1.0321
*** LIQUID PHASE ***						
RATE, LB /HR	17056.6367	0.0000	8901.9961	0.0000	14432.7441	0.0000
ACT. RATE, GBL/DAY	1250.83	0.00	787.44	0.00	1501.74	0.00
STD. LV RATE GBL/HR	51.52	0.00	27.79	0.00	44.77	0.00
CP, BTU /LB F	0.4440	0.0000	0.6245	0.0000	0.6839	0.0000
MOLECULAR WEIGHT	220.3512	0.0000	237.9579	0.0000	195.7616	0.0000
ACT. DENS, LB /FT3	56.4828	0.0000	48.3241	0.0000	41.0815	0.0000
STD. API GRAVITY	12.0000	0.0000	22.9965	0.0000	22.0322	0.0000
*** DRY BASIS ***						
RATE, LB /HR	17056.6367	0.0000	8901.9961	0.0000	14432.1133	0.0000
MOLECULAR WEIGHT	220.3513	0.0000	237.9579	0.0000	199.8498	0.0000
UOP K	10.7119	0.0000	11.1155	0.0000	10.9332	0.0000
FLASH POINT, DEG F	207.5675	0.0000	227.1025	0.0000	110.0334	0.0000
CRIT. TEMP, F	944.2889	0.0000	948.3727	0.0000	745.8058	0.0000
CRIT. PRES, PSIA	326.8333	0.0000	280.7643	0.0000	282.8885	0.0000
*** VAPOR PHASE ***						
RATE, LB /HR	0.0000	504.3504	0.0000	504.3504	15829.2266	33880.7188
ACT. RATE, FT3/SEC	0.00	1.18	0.00	0.55	4.02	7.16
STD. RATE MM FT3/DAY	0.00	0.09	0.00	0.09	0.53	0.90
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4758	1.0824	1.0249
MOLECULAR WEIGHT	0.0000	2.0174	0.0000	2.0174	11.4118	14.3287
ACT. DENS, LB /FT3	0.0000	0.1149	0.0000	0.2543	1.0927	1.3151
COMPRESSIBILITY (Z)	0.0000	1.0145	0.0000	1.0357	1.0338	1.0322
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	C.0000	0.0000
*** LIQUID PHASE ***						
RATE, LB /HR	17056.6367	0.0000	8901.9961	0.0000	14432.1133	0.0000
ACT. RATE, GBL/DAY	1250.83	0.00	787.44	0.00	1501.66	0.00
CP, BTU /LB F	0.4440	0.0000	0.6245	0.0000	C.6838	0.0000
MOLECULAR WEIGHT	220.3513	0.0000	237.9579	0.0000	199.8498	0.0000
ACT. DENS, LB /FT3	56.4828	0.0000	48.3241	0.0000	41.0818	0.0000
STD. API GRAVITY	12.0000	0.0000	22.9965	0.0000	22.0327	0.0000

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	7	3	9	11	12	13
STREAM NAME	6-202 CLTLET HOT SEP VAP	EFF FR 9-201 RECYCLE GAS	PURGE GAS	HOT SEP LIQ		
STREAM PHASE	MIXED	VAPOR	VAPOR	VAPOR	LIQUID	
TEMPERATURE, DEG F	580.3386	580.2936	418.2935	120.0000	120.0000	580.2936
PRESSURE, PSIA	1150.0000	1145.0000	1120.0000	1095.0000	1095.0000	1145.0000
RATE, LB POLS/MH	2367.8403	2345.2407	2345.2417	2140.6616	10.4499	19.5989
RATE, LB /MH	33940.1719	30240.9688	30240.9648	7476.9229	36.4994	3659.2319
ENTHALPY MM BTU /MH	14.4675	13.4662	7.6576	-3.0108	-0.0147	1.0021
ENTHALPY BTU /LO	426.2655	445.2978	253.2198	-402.6737	-402.6740	270.8992
MOLECULAR WEIGHT	14.3338	12.8751	12.5781	3.4928	3.4928	188.7464
*** VAPOR PHASE ***						
RATE, LB /MH	30215.8242	30240.9688	21046.9414	7476.9229	36.4994	0.0000
ACT.RATE, FT3/SEC	6.53	6.56	5.53	3.52	0.02	0.00
STD.RATE MM FT3/DAY	21.39	21.39	20.86	19.50	0.10	0.00
CP, BTU /LB F	1.0318	1.0300	1.1587	2.1303	2.1303	0.0000
MOLECULAR WEIGHT	12.8683	12.8781	9.1874	3.4928	3.4928	0.0000
ACT.DENS, LB /FT3	1.2548	1.2804	1.0572	0.5895	0.5895	0.0000
COMPRESSIBILITY (Z)	1.0321	1.0319	1.0331	1.0430	1.0430	0.0000
*** LIQUID PHASE ***						
RATE, LB /MH	3724.3750	0.0000	9154.0137	0.0000	0.0000	3699.2319
ACT.RATE, BBL/DAY	327.13	0.00	915.60	0.00	0.00	384.37
STD. LV RATE BBL/MH	12.09	0.00	31.34	0.00	0.00	12.00
CP, BTU /LB F	0.6748	0.0000	0.6330	0.0000	0.0000	0.6747
MOLECULAR WEIGHT	188.5288	0.0000	160.2012	0.0000	0.0000	188.7464
ACT.DENS, LB /FT3	41.1235	0.0000	42.9231	0.0000	0.0000	41.1392
STD. API GRAVITY	29.0728	0.0000	37.2206	0.0000	0.0000	29.0401
*** DRY BASIS ***						
RATE, LB /MH	3724.1621	0.0000	9193.2695	0.0000	0.0000	3699.0315
MOLECULAR WEIGHT	189.6310	0.0000	160.3039	0.0000	0.0000	188.8486
UOP K	11.2746	0.0000	11.4930	0.0000	0.0000	11.2738
FLASH POINT, DEG F	4.7693	0.0000	-40.7185	0.0000	0.0000	5.0140
CRIT. TEMP, F	709.1351	0.0000	644.0723	0.0000	0.0000	710.1290
CRIT. PRES, PSIA	290.5542	0.0000	328.0555	0.0000	0.0000	290.5071
*** VAPOR PHASE ***						
RATE, LB /MH	30155.5742	30181.7188	20528.4336	7417.4619	36.2091	0.0000
ACT.RATE, FT3/SEC	6.52	6.55	5.52	3.52	0.02	0.00
STD.RATE MM FT3/MH	0.89	0.89	0.87	0.81	0.00	0.00
CP, BTU /LB F	1.0329	1.0311	1.1607	2.1437	2.1437	0.0000
MOLECULAR WEIGHT	12.8611	12.8709	9.1749	3.4704	3.4704	0.0000
ACT.DENS, LB /FT3	1.2840	1.2796	1.0557	0.5857	0.5857	0.0000
COMPRESSIBILITY (Z)	1.0321	1.0320	1.0332	1.0430	1.0430	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE, LB /MH	3724.1621	0.0000	9193.2695	0.0000	0.0000	3659.0215
ACT.RATE, BBL/DAY	327.11	0.00	915.54	0.00	0.00	384.35
CP, BTU /LB F	0.6747	0.0000	0.6330	0.0000	0.0000	0.6747
MOLECULAR WEIGHT	189.6310	0.0000	160.3039	0.0000	0.0000	188.8486
ACT.DENS, LB /FT3	41.1234	0.0000	42.9225	0.0000	0.0000	41.1391
STD. API GRAVITY	29.0728	0.0000	37.2206	0.0000	0.0000	29.0412

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PROJECT GP JET FUELS
PROBLEM CIU204

PROCESS

SOLUTION

REFINERY PROCESSOR PROPERTIES SET

14		15	
STREAM ID.	14	15	
STREAM NAME	COLD SEP LIQ REC	COMP DIS	
STREAM PHASE	LIQUID	VAPOR	
TEMPERATURE, DEG F	120.0000	152.9113	
PRESSURE, PSIA	1095.0000	1255.0000	
RATE, LB POLS/HR	197.5457	2160.6470	
RATE, LB /HR	22735.0977	7477.0352	
ENTHALPY MM BTU /HR	0.4340	-2.4979	
ENTHALPY BTU /LB	21.2877	-334.0800	
MOLECULAR WEIGHT	115.0854	3.4923	
*** VAPOR PHASE ***			
RATE, LB /HR	0.0000	7477.0352	
ACT. RATE, FT3/SEC	0.00	3.24	
STD. RATE MM FT3/DAY	0.00	19.50	
CP, BTU /LB F	0.0000	2.1395	
MOLECULAR WEIGHT	0.0000	3.4929	
ACT. DENS, LB /FT3	0.0000	0.6413	
COMPRESSIBILITY (Z)	0.0000	1.0483	
*** LIQUID PHASE ***			
RATE, LB /HR	22735.0977	0.0000	
ACT. RATE, BBL/DAY	2167.75	0.00	
STD. LV RATE BBL/HR	85.71	0.00	
CP, BTU /LB F	0.5026	0.0000	
MOLECULAR WEIGHT	115.0854	0.0000	
ACT. DENS, LB /FT3	46.1073	0.0000	
STD. API GRAVITY	55.1029	0.0000	
*** DRY BASIS ***			
RATE, LB /HR	22727.9883	0.0000	
MOLECULAR WEIGHT	115.2766	0.0000	
UOP K	11.9576	0.0000	
FLASH POINT, DEG F	-91.1659	0.0000	
CRIT. TEMP, F	523.3317	0.0000	
CRIT. PRES, PSIA	404.9004	0.0000	
*** VAPOR PHASE ***			
RATE, LB /HR	0.0000	7417.5693	
ACT. RATE, FT3/SEC	0.00	3.23	
STD. RATE MM FT3/DAY	0.00	0.81	
CP, BTU /LB F	0.0000	2.1530	
MOLECULAR WEIGHT	0.0000	3.4704	
ACT. DENS, LB /FT3	0.0000	0.6371	
COMPRESSIBILITY (Z)	0.0000	1.0484	
VISCOSITY, CP	0.0000	0.0000	
*** LIQUID PHASE ***			
RATE, LB /HR	22727.9883	0.0000	
ACT. RATE, BBL/DAY	2167.75	0.00	
STD. LV RATE BBL/HR	85.71	0.00	
CP, BTU /LB F	0.5026	0.0000	
MOLECULAR WEIGHT	115.2766	0.0000	
ACT. DENS, LB /FT3	46.1073	0.0000	
STD. API GRAVITY	55.1029	0.0000	

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SIMULATION SCIENCES INC.
PROJECT GP JET FUELS
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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	25	26	27	28	30	31
STREAM NAME	HDC NAPHTHA	HDC JP-4	FRAC CVMD	HDC HC VAP	FRAC STR STM JP-4	STR STM
STREAM PHASE	LIQUID	LIQUID	VAPOR	VAPOR	VAPOR	VAPOR
TEMPERATURE, DEG F	90.0000	262.8516	200.4364	90.0000	322.6034	312.2666
PRESSURE, PSIA	60.0000	55.3000	70.0000	60.0000	74.7998	55.3000
RATE, LB POLS/MR	35.9516	115.9792	425.3399	25.6295	33.3056	37.1912
RATE, LB /MR	2853.6284	14037.0020	25687.5742	706.5347	599.9996	609.9996
ENTHALPY MM BTU /MR	0.0515	1.3050	6.6696	0.0994	0.7141	0.7974
ENTHALPY BTU /LB	18.0337	92.9704	259.6436	140.7007	1190.2109	1190.2109
MOLECULAR WEIGHT....	71.4272	121.0303	60.3930	27.5672	18.0150	18.0150
*** VAPOR PHASE ***						
RATE, LB /MR	0.0000	0.0000	25687.5742	706.5347	599.9996	609.9996
ACT. RATE, FT3/SEC	0.00	0.00	11.14	0.69	1.00	1.00
STD. RATE MM FT3/DAY	0.00	0.00	3.87	0.23	0.30	0.36
CP, BTU /LB F	0.0000	0.0000	0.5009	0.5283	0.5550	0.5307
MOLECULAR WEIGHT....	0.0000	0.0000	60.3930	27.5672	18.0150	18.0150
ACT. DENS, LB /FT3	0.0000	0.0000	0.6403	0.2855	0.1674	0.1242
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9321	0.9820	0.9589	0.9685
*** LIQUID PHASE ***						
RATE, LB /MR	2853.6284	14037.0020	0.0000	0.0000	0.0000	0.0000
ACT. RATE, BBL/DAY	311.46	1443.16	0.00	0.00	0.00	0.00
STD. LV RATE BBL/MR	12.64	52.79	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.5444	0.5854	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	71.4272	121.0303	0.0000	0.0000	0.0000	0.0000
ACT. DENS, LB /FT3	39.1644	41.5768	0.0000	0.0000	0.0000	0.0000
STD. API GRAVITY....	87.7928	54.6287	0.0000	0.0000	0.0000	0.0000
*** DRY BASIS ***						
RATE, LB /MR	2352.8999	13990.8379	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	71.4813	123.3578	0.0000	0.0000	0.0000	0.0000
UOP K	12.7357	11.9769	0.0000	0.0000	0.0000	0.0000
FLASH POINT, DEG F	-86.2333	33.1789	0.0000	0.0000	0.0000	0.0000
CRIT. TEMP, F	383.9533	600.5282	0.0000	0.0000	0.0000	0.0000
CRIT. PRES, PSIA	503.1559	397.5057	0.0000	0.0000	0.0000	0.0000
*** VAPOR PHASE ***						
RATE, LB /MR	0.0000	0.0000	24472.5000	701.1978	0.0000	0.0000
ACT. RATE, FT3/SEC	0.00	0.00	9.27	0.68	0.00	0.00
STD. RATE MM FT3/MR	0.00	0.00	0.14	0.01	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.5031	0.5290	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	60.3930	27.5672	0.0000	0.0000
ACT. DENS, LB /FT3	0.0000	0.0000	0.7332	0.2868	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9217	0.9818	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0091	0.0108	0.0000	0.0000
*** LIQUID PHASE ***						
RATE, LB /MR	2552.8999	13990.8379	0.0000	0.0000	0.0000	0.0000
ACT. RATE, BBL/DAY	311.41	1434.79	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.5443	0.5955	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	71.4813	123.3578	0.0000	0.0000	0.0000	0.0000
ACT. DENS, LB /FT3	39.1607	41.5373	0.0000	0.0000	0.0000	0.0000
STD. API GRAVITY....	87.7928	54.6287	0.0000	0.0000	0.0000	0.0000

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.
STREAM NAME
STREAM PHASE
TEMPERATURE, DEG F
PRESSURE, LB /MR
RATE, LB POLS/MR
RATE, LB /MR
ENTHALPY MM BTU /MR
ENTHALPY BTU /LB
MOLECULAR WEIGHT....
SOUR H2O
LIQUID
90.0000
60.0000
66.8175
1203.7168
6.0698
58.0156
19.0150
90.0000
60.0000
292.513
20923.9305
0.3773
19.0337
71.272

*** VAPOR PHASE ***
RATE, LB /MR
ACT. RATE, FT3/SEC
STD. RATE MM FT3/DAY
CP, BTU /LB F
MOLECULAR WEIGHT....
ACT. DENS, LB /FT3
COMPRESSIBILITY (Z).
0.0000
0.00
0.00
0.0000
0.0000
0.0000
0.0000
0.0000
0.0000

*** LIQUID PHASE ***
RATE, LB /MR
ACT. RATE, OBL/DAY
STD. LV RATE OBL/MR
CP, BTU /LB F
MOLECULAR WEIGHT....
ACT. DENS, LB /FT3
STD. API GRAVITY....
1203.7168
82.86
3.44
0.9978
18.0150
62.1002
10.0635
20923.9305
2293.73
92.71
0.3644
71.4272
39.1644
87.7927

*** DRY BASIS ***
RATE, LB /MR
MOLECULAR WEIGHT....
UOP K
FLASH POINT, DEG F
CRIT. TEMP, F
CRIT. PRES, PSIA
20918.6406
71.4814
12.7357
-96.2333
383.9538
503.1559

*** VAPOR PHASE ***
RATE, LB /MR
ACT. RATE, FT3/SEC
STD. RATE MM FT3/MR
CP, BTU /LB F
MOLECULAR WEIGHT....
ACT. DENS, LB /FT3
COMPRESSIBILITY (Z).
VISCOSITY, CP
0.0000
0.00
0.00
0.0000
0.0000
0.0000
0.0000
0.0000
0.0000

*** LIQUID PHASE ***
RATE, LB /MR
ACT. RATE, OBL/DAY
CP, BTU /LB F
MOLECULAR WEIGHT....
ACT. DENS, LB /FT3
STD. API GRAVITY....
20913.6406
2293.36
0.3643
71.414
39.1647
87.7927

VERSION 2
SIMULATION SCIENCES INC.
PROJECT GP JET FUELS
PROBLEM STABILIZER

TM
PROCESS
SOLUTION

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.....	29	40	41	42	43	44
STREAM NAME.....	HOT NAP	NAP FEED	STAB OFFGAS	REFLUX	NAP PROD	REBOILER
STREAM PHASE.....	LIQUID	LIQUID	VAPOR	LIQUID	LIQUID	LIQUID
TEMPERATURE, DEG F	90.0000	89.8248	94.1650	94.1650	262.5658	226.5963
PRESSURE, PSIA	150.0000	144.0000	125.0000	125.0000	142.0000	141.5000
RATE LB ROLS/MR	44.8024	84.7363	7.1663	7.9156	77.5554	146.9895
RATE LB /MR	3209.6333	6063.4775	294.8259	951.0134	5767.9932	10470.0488
ENTHALPY MM BTU /MR	0.0046	0.0001	0.0616	0.0347	0.6437	0.9397
ENTHALPY BTU /LB	2.6839	9.9054	209.0047	36.4371	111.9418	91.6606
MOLECULAR WEIGHT....	71.6398	71.5397	41.1406	55.0829	74.3725	71.2299
*** VAPOR PHASE ***						
RATE LB /MR	0.0000	0.0000	294.8259	0.0000	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.08	0.00	0.00	0.00
STD.RATE MM FT3/DAY	0.00	0.00	0.07	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.4772	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	41.1406	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.9812	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9818	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /MR	3209.6333	6063.4775	0.0000	951.0134	5767.9932	10470.0488
ACT.RATE BBL/DAY	318.46	629.37	0.00	122.78	716.00	1275.42
STD. LV RATE BBL/MR	12.93	25.62	0.00	4.88	24.03	44.54
CP, BTU /LB F	0.5048	0.5233	0.0000	0.7108	0.6893	0.6920
MOLECULAR WEIGHT....	71.6398	71.5397	0.0000	53.0829	74.3725	71.2299
ACT.DENS LB /FT3	43.0820	41.1822	0.0000	33.1094	34.6351	35.0904
STD. API GRAVITY....	68.6129	77.6396	0.0000	122.6633	74.8996	79.0423
*** DRY BASIS ***						
RATE LB /MR	3208.8169	6061.9326	0.0000	950.6528	5767.8945	10468.9688
MOLECULAR WEIGHT....	71.6941	71.5399	0.0000	53.1222	74.3765	71.2516
UOP K	11.7920	12.2361	0.0000	13.8836	12.1437	12.2821
FLASH POINT, DEG F	-73.3390	-83.7349	0.0000	-118.4625	-57.6394	-77.2691
CRIT. TEMP, F	722.6771	404.6239	0.0000	263.5591	421.0440	401.0717
CRIT. PRES, PSIA	575.3867	541.3368	0.0000	603.4601	515.4408	529.5265
*** VAPOR PHASE ***						
RATE LB /MR	0.0000	0.0000	294.0104	0.0000	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.08	0.00	0.00	0.00
STD.RATE MM FT3/MR	0.00	0.00	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.4775	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	41.2876	0.0000	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.9856	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9811	0.0000	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0090	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /MR	3208.8169	6061.9326	0.0000	950.6528	5767.8945	10468.9688
ACT.RATE BBL/DAY	318.40	629.26	0.00	122.76	716.00	1275.34
CP, BTU /LB F	0.5047	0.5232	0.0000	0.7107	0.6893	0.6920
MOLECULAR WEIGHT....	71.6941	71.5399	0.0000	53.1222	74.3765	71.2516
ACT.DENS LB /FT3	43.0727	41.1737	0.0000	33.1036	34.6349	35.0889
STD. API GRAVITY....	68.6123	77.6392	0.0000	122.7061	74.9007	79.0494
VISCOSITY, CP	0.0000	0.0000	0.0090	0.0000	0.0000	0.0000

APPENDIX C

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 2
PROFITABLE JP-4 PRODUCTION
SUBTASKS 1.2 & 1.3
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
DATE - JAN. 30, 1988

W16242/bf

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1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to produce JP-4 type aviation turbine fuel and chemical byproducts to maximize profit from Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- . Tar Oil byproduct stream (47620 #/hr, 3182 BPSD) is charged to the hydrotreater (Area 100).
- . The hydrotreater is a 3 stage expanded bed type process which removes 99% + of the sulfur, nitrogen, and oxygen compounds and begins the conversion of 500°F+ material. The hydrotreater adds a large quantity of hydrogen to the feed (3400 SCF/bbl) which results in a high heat of reaction. An expanded bed type reactor was chosen to both control and utilize the heat of reaction. Three stages were used to both control the temperature rise as well as to obtain the high efficiency associated with staging a back-mixed reactor.
- . The hydrotreater produces 6 streams:
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H₂ and CH₄.
 - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.
 - Unstabilized naphtha which is sent to the combined naphtha stabilizer in the hydrocracker (area 200). After stabilization, to control vapor pressure, the naphtha is sent to storage and gasoline blending.
 - JP-4 turbine fuel which is combined with JP-4 produced in the hydrocracker (area 200) and sent to storage.
 - 500°F+ unconverted bottoms product which is sent to the hydrocracker (area 200).
 - Wastewater containing, NH₄OH and NH₄HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H₂S and NH₃.

- Approximately 950 #/day of spent catalyst which is shipped to a catalyst reclaimer in the same drums that the catalyst is received in.
- . The 500⁰F+ unconverted stream from the expanded bed hydrotreater (area 100) is charged to the fixed bed hydrocracker (area 200). The hydrocracker converts this material to naphtha and JP-4 turbine fuel. For this service a 5 stage unit was chosen with 65% conversion per pass. This unit also includes a naphtha stabilizer which stabilizes both the naphtha produced in the hydrotreater and hydrocracker.
- . The hydrocracker produces 4 streams in addition to JP-4
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit of the SNG plant for recovery of the H₂ and CH₄.
 - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.
 - Stabilized naphtha which is sent to storage and gasoline blending.
 - A small sour water stream which is sent to the PHOSAM unit in the SNG plant or alternatively used as part of the injection water to the hydrotreating plant.
- . Hydrogen make-up for the Hydrotreater, the Hydrocracker and the Naphtha Hydrotreater (Area 600) is supplied from a PSA Hydrogen Unit (Area 300). High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining purge gas is available a low pressure (5 psig) which has a fuel value of about 565 BTU/ft³. This H₂, CO & CH₄ rich gas is recompressed into the methanation unit of the SNG plant.
- . The crude naphtha byproduct stream (8738#/hr, 725 BPSD) is charged to the distillation and hydrotreating unit (Area 600).
- . The distillation removes the material boiling below 160⁰F, which is sent to the SNG Plant fuel pool, and produces a bottoms product which is charged to the hydrotreater.
- . The fixed bed hydrotreater is a single bed reactor which removes 99% + of the sulfur, nitrogen, and oxygen compounds. Hydrogen is added to the feed at the rate of 430 SCF/bbl.

1.1 Overall Process Description - cont'd

- . The naphtha hydrotreater produces 4 streams:
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4 .
 - Naphtha which is stabilized to control vapor pressure. Approximately 74% of the naphtha is sent to the Aromatics Recovery Unit (Area 700), the remainder is sent to gasoline blending.
 - A low pressure off gas which is sent to the Stretford unit in the SNG plant.
 - Wastewater containing, NH_4OH and NH_4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H_2S and NH_3 .
- . A portion of the hydrotreated naphtha is charged to the extraction section of the Aromatics Recovery Unit (Area 700) where it is contacted with a solvent to extract the aromatic components from the stream. The raffinate is sent to storage and gasoline blending while the solvent is recovered from the aromatic extract. The aromatic extract is then sent to fractionation to produce the BTX products.
- . Five streams are produced in the ARU plant.
 - A hydrocarbon gasoline blending stock which is sent to storage and gasoline blending.
 - A small process water stream which is sent to the waste treatment plant in the SNG Plant.
 - Three product streams Benzene, Toluene & Xylene which are sent to storage.
- . The crude phenol byproduct stream (14490 #/hr, 936 BPSD), is feed to the dual solvent phenol extraction unit (Area 800).

1.1 Overall Process Description - cont'd

- . Distillation removes approx. 85% of the phenol which is further distilled to remove light ends and then reflashed over sulfuric acid producing a 99.8% pure product.
- . The remainder of the stream (a cresylic acid mixture) is flash distilled over a 3 wt.% concentrated sulfuric acid mixture to remove pyridine type substances.
- . The acid tar produced is water washed and mixed with light oil and sent to fuel.
- . The remaining cresol/xylene mixture is double solvent extracted to remove neutral hydrocarbons. The resulting crude cresylic acid is dried and sent either to storage or distillation (Area 900).
- . Streams produced in the phenol extraction unit are:
 - Phenol product sent to storage
 - Crude Cresylic Acid sent to distillation (Area 900) or storage.
 - Wash Water sent to Water Treatment in the SNG Plant.
 - Waste Water sent to the Phenosolvan unit in the SNG Plant.
 - Neutral Oil sent to storage and fuel for the SNG Plant boilers.
- . The Crude Cresylic Acid is progressively distilled (Area 900) to separate the cresols and xylenols. No attempt has been made to remove the guaiacol from the product streams.
- . Streams produced in the crude cresylic acid distillation unit are:
 - o-Cresol product which is sent to storage.
 - m,p-Cresol product which is sent to storage.
 - Xylene product which is sent to storage.
 - A heavy distillate which is combined with neutral oil in Area 800.

1.1 Overall Process Description - cont'd

- A crude phenol stream which is recycled to the Area 800.
- A small water stream which is sent to Area 800 for tar acid washing.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas, neutral oil and 160°F-distillate produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

936 BPSD of Crude Phenol
725 BPSD of Crude Naphtha
3182 BPSD of Tar Oil
4347 BPSD of #6 Fuel Oil
9.52 MMSCFD equivalent SNG product loss due to the syn gas feed to the PSA unit.

Products

3403 BPSD of JP-4 turbine fuel
324 BPSD of 160°F - Naphtha for gasoline blending
317 BPSD of Phenol
56 BPSD of o-Cresol
131 BPSD of m,p-Cresol
75 BPSD of Xylenols
312 BPSD of Neutral Oil for Fuel
202 BPSD of 160°F - Distillate for Fuel
161 BPSD of Gasoline Blending Stock
233 BPSD of Benzene
83 BPSD of Toluene
11 BPSD of Xylene
6.27 MMSCFD equivalent SNG product credit due to HDT, HDC & PSA purge gas reinjection into SNG plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	4347 BPSD
SNG Equivalent of Syn Gas & Purge Gas	3.25 MM SCFD
Power	6180 kW
Cooling Water	6140 GPM (30°F rise)
Process Water	31.0 GPM

In addition the process utilizes steam as summarized below which was debited against boiler requirements.

HP Steam Import	54,700 #/H
MP Steam Import	8,900 #/H
LP Steam Export	6,900 #/H
Condensate Return	63,600 #/H

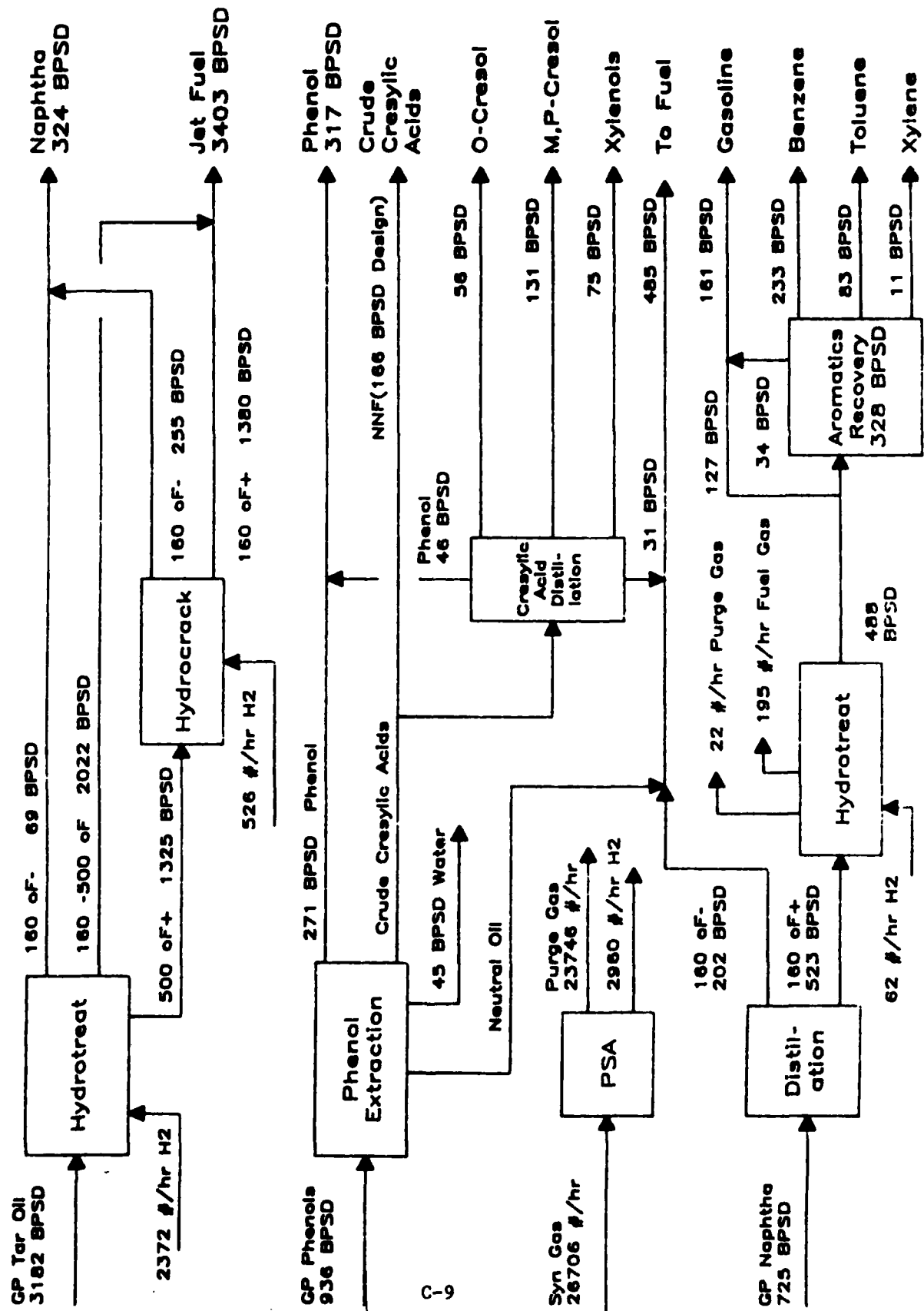


Table 1.1 Great Plains Case 2: Profitable JP4 Production

=====					
Tar Oil Feed====>	47620	#/hr	3182	BPSD	
Phenol Feed====>	14490	#/hr	936	BPSD	
Crd Naphtha Feed=>	8738	#/hr	725	BPSD	
Naphtha Product==>	3370	#/hr	324	BPSD	
JP4 Product=====>	40117	#/hr	3403	BPSD	
Phenol Product====>	4925	#/hr	317	BPSD	
o-Cresol Prod====>	845	#/hr	56	BPSD	
m,p-Cresol Prod====>	1974	#/hr	131	BPSD	
Xylenols Prod====>	1070	#/hr	75	BPSD	
Gasoline Stock====>	1957	#/hr	161	BPSD	
Benzene Prod=====>	3003	#/hr	233	BPSD	
Toluene Prod=====>	1054	#/hr	83	BPSD	
Xylene Prod=====>	139	#/hr	11	BPSD	
SN6 Product Loss=>	5586	#/hr	3.3	MMSCFD	
Fuel Oil Makeup==>	60183	#/hr	4347	BPSD	

Expanded Bed Hydrotreater

=====					
Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD

Feeds					
H2	4.98		2373	1176.9	
Tar Oil	100.00	1.0268	47620		3182

Total	104.98		49993		
Products					
Purge Gas	0.11		54	16.8	
Fuel Gas	1.99		950	46.8	
Naphtha	1.61	0.6825	765		77
JP-4	50.86	0.8218	24221		2022
500 oF+	39.95	0.9850	19026		1325
H2O in SW	8.96		4267	237.1	
H2S in SW	0.43		205	6.0	
NH3 in SW	1.06		505	29.7	

Total	104.98		49993		3424

Fixed Bed Hydrocracker

=====					
Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD

Feeds					
H2	2.76		526	260.9	
500oF+	100.00	0.9850	19026		1325

Total	102.76		19552		
Products					
Purge Gas	0.94		179	60.7	
Fuel Gas	4.31		820	45.3	
Naphtha	13.96	0.6675	2656		273
JP-4	83.55	0.7900	15896		1381
H2S in SW	0.003		0.6	0.02	
NH3 in SW	0.003		0.6	0.03	

Total	102.77		19552		1654

Naphtha Stabilizer

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
HDT Nap	22.37	0.6825	765		77
HCR Nap	77.63	0.6675	2656		273
Stab Nap	98.50	0.7140	3370		324
Fuel Gas	1.50		51	1.2	

PSA Hydrogen Recovery Unit(86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	8.63	5.55	903.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	1707.6	503.1	39.9	438.2	8.5	5.1	2702.6
Prod. H2	1468.5	0.1	0.0	0.0	0.0	0.0	1468.7
Purge Gas	239.1	503.0	39.9	438.2	8.5	5.1	1233.9
#/hr							
Feed Gas	3443	14091	1760	7030	255	164	26743
Prod. H2	2961	4	0	0	0	0	2964
Purge Gas	482	14087	1760	7030	255	164	23779

Crude Naphtha
Distillation

	Wt %	Gravity	#/hr	BPSD
Feed Naphtha	100.00	0.8269	8738	725
Prod 160 oF-	24.77	0.7350	2164	202
Prod 160 oF+	75.23	0.8627	6574	523

Naphtha Hydrotrater

=====

Component	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feed 160 oF+	100.00	0.8627	6574		523
Feed Hydrogen	0.94		62	30.8	
Feed Total	100.94		6636		523
Products					
Purge Gas	0.33		22	6.8	
Fuel Gas	2.97		195	10.8	
HDT Naphtha	93.61	0.8650	6154		488
H2O in SW	1.96		129		
H2S in SW	1.76		116		
NH3 in SW	0.30		20		
Total Products	100.94		6636		488

Aromatics Recovery

=====

Component	Wt %	Grav	#/hr	BPSD
Feed HDT Naphtha	100.00	0.8650	4552	361
Products				
Raffinate	7.82	0.7175	356	34.0
Benzene	65.97	0.8844	3003	233.0
Toluene	23.15	0.8718	1054	83.0
Xylene	3.05	0.8729	139	10.9
Total Products	100		4552	361

Phenol Extraction

=====

Component	Wt %	#/hr	Grav	BPSD
Feeds				
Crude Phenol	100.00	14490	1.0621	936
Sulfuric Acid	1.97	285	1.8300	11
Total Feed	101.97	14775		947
Products				
Cr. Cresylic Acid	35.13	5090	1.0290	339
Phenol	29.09	4215	1.0661	271
Neutral Oil	30.68	4445	1.0860	281
Acidic Waste Water	7.07	1025	1.2558	56
Total Products	101.97	14775		947

Cresylic Acid Distillation

Component	Wt %	#/hr	Grav	BPSD
Cr. Cresylic Acid	100.00	5090	1.0290	339
Products				
Phenol	13.95	710	1.0661	46
o-Cresol	16.60	845	1.0350	56
m,p-Cresol	38.78	1974	1.0340	131
Xylenols	21.02	1070	0.9750	75
Heavies	9.65	491	1.0800	31
Total	100.00	5090		339

Fuel Gas Generated in Hydrotreating and Hydrocracking

Component	#/hr	#Mol/hr	MMBTU/hr
HDTR FG Produced	950	46.8	17.1
HCR FG Produced	820	45.3	14.8
Stabilizer FG	51	1.2	0.9
Naphtha Hdtr FG	195	10.8	3.5
Total Fuel Gas	1821	93.3	32.8

Fuel Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	482	239.1	324	29.4
CO	14087	503.0	321	61.2
CO2	1760	39.9	0	0.0
C1	7030	438.2	1010	167.7
C2	255	8.5	1769	5.7
N2+Ar	164	5.1	0	0.0
Total	23779	1233.9	565	264.0

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft ³	BPSD
Tar Oil	-47620	17000	-809.5			-3182
Crude Phenol	-14490	13070	-189.4			-936
Crude Naphtha	-8738	18500	-161.7			-725
Fuel Gas	1821	18000	32.8	0.8	927	
160 oF- distillate	2164	17400	37.7			202
Neutral Oil	4936	15000	74.0			312
Import Steam	-56700	1000	-56.7			
Fuel Oil to Boiler	59600	18000	1072.8			4305
Total	-59027		0.0	0.8		-24
Fuel Oil to Process Heaters	583	18000	10.5			42

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	9.52	64862
SNG Credit for PSA Purge gas	5.96	29613
SNG Credit for Hdtrs purge gas	0.31	2023
Total SNG Production Loss	3.25	

2.0 PROCESS DESCRIPTION

2.1 Hydrotreater (Area 100)

For a description of the Hydrotreater process see Case 3 Section 2.1.

2.2 Hydrocracker (Area 200)

For a description of the Hydrocracker process see Case 1 Section 2.2.

2.3 PSA Hydrogen Unit & Recompression (Area 300)

2.3.1 Hydrogen for both hydrotreaters and the hydrocracker will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig
Temp.	65 °F
Composition	mol%
H2	63.19
CO	18.61
CO2	1.48
CH4	16.21
C2H6	0.31
COS, H2S, CS2	< 0.01
N2 + Ar	0.19
H2O	< 0.01

The PSA unit selectively absorbs all components except H2 and produces a 99.99% vol. purity stream at about 345 psig and 80°F. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100°F
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

At the conditions given a 10 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

2.3 PSA Hydrogen Unit & Recompression (Area 300) - cont'd

2.3.1 Cont'd

The system uses 10 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continuously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-20301 presents a schematic of a Union Carbide Polybed PSA unit.

2.3.2 The purge gas is recompressed to 375 psia and sent to the methanation unit of the SNG plant.

2.4 Phenol Stream (Area 800 and 900)

For description of the Phenol Extraction (Area 800) and Cresylic Acid Distillation (Area 900) Units see Case 7 Section 2.1.

2.5 Naphtha Stream (Areas 600 and 700)

For a description of the Naphtha Distillation and Hydrotreating Unit (Area 600) and the Aromatics Recovery Unit (Area 700) see Case 7 Section 2.2.

The diagram illustrates a hydrogen gas distribution system. A main horizontal line at the top, labeled 'HYDROGEN' with an arrow pointing right, contains 10 valve stations numbered 20 to 29. Below this line, a series of vertical and horizontal pipes connect to 10 cylindrical 'ADSORBERS' numbered 1 to 10. Each adsorber has two outlet lines: one leading to a lower horizontal line with valve stations 40 to 49, and another leading to a bottom horizontal line with valve stations 30 to 39. The bottom horizontal line is labeled 'PURGE GAS STREAM' with an arrow pointing right. A central vertical line with valve stations 60 to 65 serves as a manifold for the purge gas stream. A circled number '1' is located at the bottom right corner of the diagram.

PAGE 2-3

TYPICAL ARRANGEMENT

NUMBER OF ABSORBERS FOR THIS CASE=10

ARRANGEMENT	3:30RBERS FOR THIS CASE = 10									
△										
△	W	FOR SUBTASK 1-2	ML	E.S.						
REV	DATE	DESCRIPTION	PROG	PROG	PROG	PROG	PROG	PROG	PROG	PROG
THE LUMBUS COMPANY		LUMBUS								
TITLE		PSA HYDROGEN UNIT								
CLIENT		AMOCO/BOE								
PROJECT NO		5571								
CASE		2								
PROJECT NO		5571-20301								

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.1 Equipment List

CASE 2 - PROFITABLE JP-4

AREA 100 - HYDROTREATER

TAG. NO. DESCRIPTION

See Case 3 Area 100

AREA 200 - HYDROCRACKER

See Case 1 Area 200

AREA 300 - PSA HYDROGEN UNIT & RECOMPRESSION

FA-301 Purge Gas Surge Drum

GB-301 Purge Gas Compressor

PA-301 PSA Hydrogen Unit Package

AREA 400 - STORAGE AREA

FB-401 Jet Fuel Storage Tank

FB-402 Naphtha Storage Tank

FB-403 Fuel Oil Storage Tank

FB-404 Blending Stock

FB-405 Benzene Storage

FB-406 Toluene Storage

FB-407 Xylene Storage

FB-409 Gasoline Storage

FB-410 Neutral Oil Storage

FB-411 Phenol Product Storage

FB-412 Crude Cresylic Acid Storage

FB-413 O-Cresol Storage

FB-414 M, P Cresol Storage

FB-415 Xylenol Storage

3.0 CAPITAL COSTS

3.1 Equipment List - cont'd

CASE 2 - PROFITABLE JP-4

<u>TAG NO.</u>	<u>DESCRIPTION</u>
<u>AREA 400</u>	<u>STORAGE AREA</u>
GA-401A/S	Tar/Tar Oil Feed Pump
GA-402A/S	Crude Phenol Feed Pump
GA-403A/S	Fuel Oil Transfer Pump
GA-404A/S	Naphtha Transfer Pump
GA-405A/S	Crude Naphtha Transfer Pump
GA-406A/S	Blending Stock Pump
GA-407A/S	Benzene Transfer Pump
GA-408A/S	Toluene Transfer Pump
GA-409A/S	Xylene Transfer Pump
GA-411A/S	Gasoline Transfer Pump
GA-412A/S	Neutral Oil Transfer Pump
GA-413A/S	Crude Cresylic Acid Transfer Pump
GA-414A/S	O-Cresol Transfer Pump
GA-415A/S	M, P. Cresol Transfer Pump
GA-416A/S	Xylenol Transfer Pump

PA-401 Gasoline Blending Package

AREA 500 - CATALYST HANDLING

See Case 3 Area 500

AREA 600 - NAPHTHA DISTILL. AND HDT

See Case 7 Area 600

AREA 700 - AROMATICS RECOVERY

See Case 7 Area 700

AREA 800 - PHENOL EXTRACTION

See Case 7 Area 800

AREA 900 - CRESYLIC ACID DISTILLATION

See Case 7 Area 900

3.2 Cost Estimate

3.2.1 Basis of Estimate

The estimate for this case is a factored type estimate using the T.I.C. values developed for the various cases referenced in this project.

The total investment costs are scaled to the capacity requirement of this case using a 0.6 exponent.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs for Areas 100 thru 700 and 30% for Areas 800 & 900.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2 Estimate Summary

(Thousands of \$)

	<u>Case 2</u>
Area 100 Hydrotreater	\$20,702
Area 200 Hydrocracker	10,430
Area 300 PSA & Recompression	8,100
Area 400 OSBL	8,443
Area 500 Catalyst Handling	1,285
Area 600 Naph. Dist & HDT	4,615
Area 700 ARU	7,887
Area 800 Phenol Ext.	12,276
Area 900 Cresylic Acid Dist.	4,832
Subtotal	\$78,570
Area 700 ARU Solvent Inventory	80
Total	<u>\$78,650</u>

3.2.3 Estimate Breakdown (Area 100) All Values in Thousands

This unit has the same capacity as the 100 Area of Case 3.
Therefore, T.I.C. = \$20,702.

Area 200

This unit has a 107% capacity of the 200 Area of Case 1.
Therefore, T.I.C. = $(1.07)^{0.6} (10,012) = \$10,430$

3.2.3 Estimate Breakdown - Cont'd

Area 300

This unit has a 96% capacity of the 300 Area of Case 3.
Therefore T.I.C. = $(0.96)^{0.6} (8300) = \$8100$.

Area 400

Tar Oil Stream Storage 80% Case 1
TIC = $(0.8)^{0.6} (5 \text{ m}) = \$4300$

Phenol Stream Storage = 100% Case 7
TIC = \$3,016

Naphtha Stream Storage 75% Case 7
TIC = $(0.75)^{0.6} (3058) = \$2,890$
Subtotal 10,206

Less Duplicate Pipe & Rack -1763
Total = \$8,443

Area 500

The capacity of this unit is identical to the 500 Area of Case 3. Therefore T.I.C. = \$1,285

Area 600

This unit has a capacity of 100% of the 600 Area of Case 7.
Therefore T.I.C. = \$4,615

Area 700

This unit has a capacity of 75% of the 700 Area of Case 7.
Therefore T.I.C. = $(0.75)^{0.6} (9,373) = \$7,887$

Area 800

This unit has a capacity of 100% of the 800 Area of Case 7.
Therefore T.I.C. = \$ 12,276

Area 900

This unit has a capacity of 100% of the 900 Area of Case 7.
Therefore = T.I.C. = \$ 4,832

4.0 OPERATING COSTS

4.1 Operating Labor

It is estimated that it will require men/shift to operate the plant broken down as follows:

Foreman	2
Control Room	2
HDT Operator	2
HCR Operator	2
PSA & relief man	1
Naphtha Distil. & HDT	2
ARU	2
Phenol Ext.	1
Cresylic Acid Dist.	<u>1</u>
	15 Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 7 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	15 positions x 4 people/position -	60
Supervisor & Admin.		6
QC Technician		2
Maintenance		7
Other (Stores or Janitorial)		<u>1</u>
Total		76

4.2 Utilities

The following utilities have been estimated:

<u>Utility</u>	<u>Consumption</u>	<u>Cost</u>	<u>\$/SD</u>
#6 Fuel Oil	4347 BPSD	\$16/Bbl (a)	69,552
SNG equivalent	3.25 MMSCFD	\$3.80/MM Btu (b)	12,105
of Syn Gas & Purge Gas			
Cooling Water	6140 GPM	\$0.155/MGal (c)	1,370
Power	6180 kW	\$0.04/kWH (c)	5,933
Process Water	31 GPM	\$0.45/MGal (c)	20

(a) Cut of 1% sulfur \$6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

4.2 Utilities - cont'd

(b) Memo from D. Daley of Burns & Roe to L. Lorenzo of DOE dated Oct. 20, 1987, reference DPD-87-863.

(c) ANG utility cost information dated 5/87.

4.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

<u>Catalyst & Chem.</u>	<u>Use</u>	<u>Cost</u>	<u>\$/SD</u>
Nap. HDT Cat	0.021 #/Bbl	\$3.00/#	33
HDT Cat.	0.30 #/Bbl	\$3.00/#	2864
HCR Cat.	0.013 #/Bbl	\$6.00/#	104
Inhibitors	50 PPM	\$10/Gal	92
ARU Solvent	18 #/D	\$2.10/#	38
H ₂ SO ₄	7100 #/D	\$0.04/#	285
			<u>\$3416</u>

4.4 Maintenance Supplies

Maintenance supplies for hydrotreating operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units (excluding ARU solvent inventory). On this basis the maintenance supplies would be

$$0.00005 \times 78,570,000 = \$ 3929/SD$$

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the production of JP-4 and by-product chemicals are proposed to be located to the east of the Phenosolvan Unit and Water Treatment Area of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 340' x 575' will be surrounded by an access road and will be divided by three central east-west roads. Areas 100 & 500 will be located to the north and Areas 200 & 300 South of Area 100, Areas 800 & 900 next and then Areas 600 & 700.

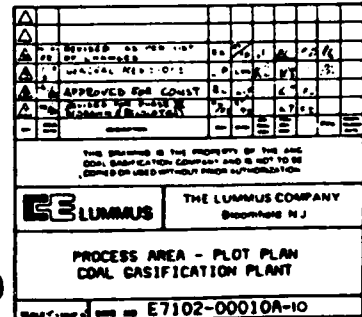
A diked storage tank area approx. 375' x 375' will be required for product and fuel oil storage and is proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

5.2 Unit Tie-Ins

Approximately 2500 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

A summary of the lines has not been prepared for this case but will be similar to a combination of Cases 3 and 7 with the utility lines of like services being combined.



CASE 2

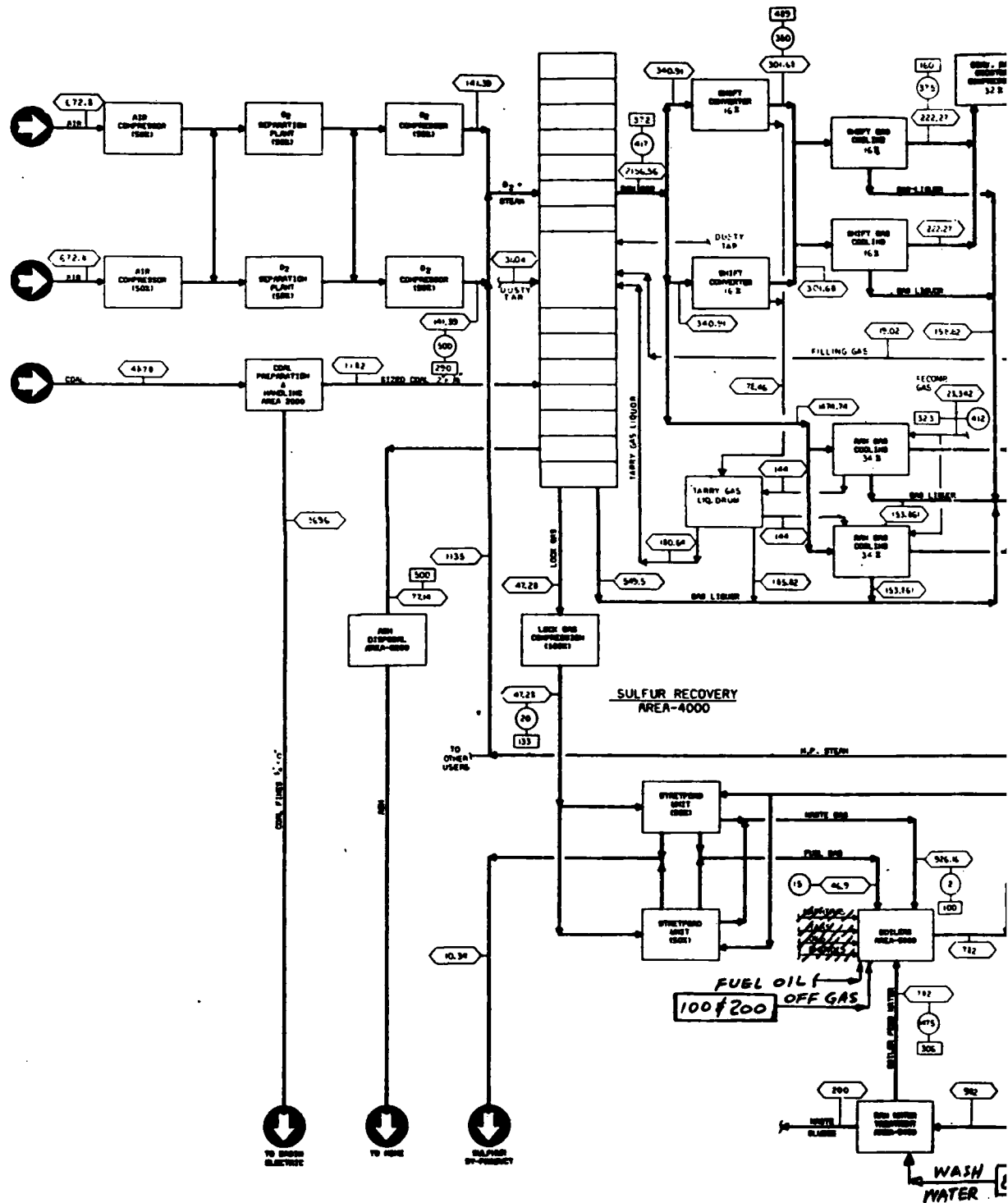
**COAL PREPARATION
AND HANDLING
AREA-2000**

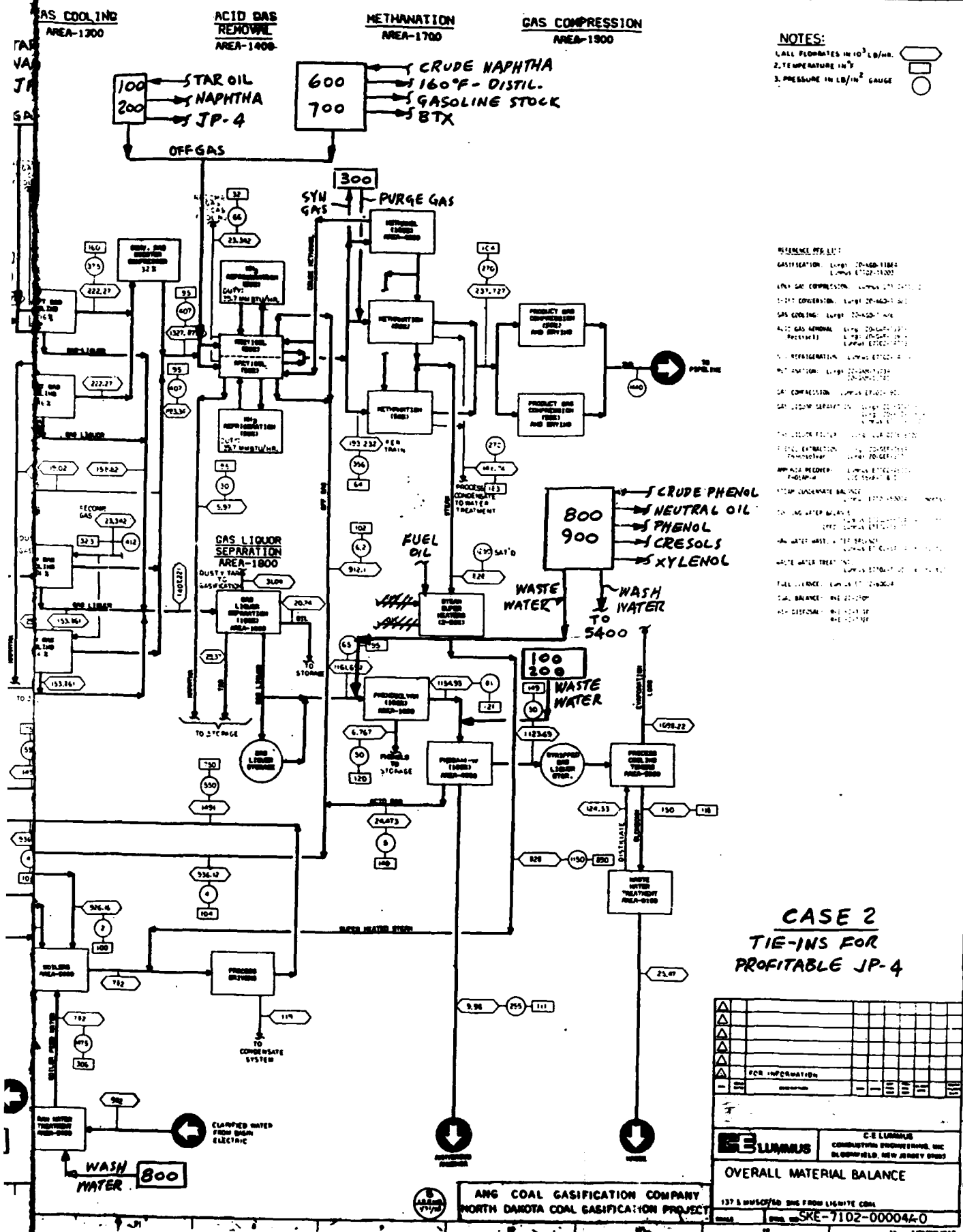
**OXYGEN PLANTS
AREA-3000**

**GASIFICATION
AREA-1100**

**SHIFT CONVERSION
AREA-1200**

**GAS COOLING
AREA-1300**





NOTES:

1. ALL FLOW RATES IN 10^3 LB/HR.
2. TEMPERATURE IN $^{\circ}$ F
3. PRESSURE IN LB/IN ^{2} GAUGE

DISPATCH NO. 41:

[illegible]

CASE 2
TIE-INS FOR
PROFITABLE JP-4

[illegible]

APPENDIX D

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 3
MAXIMUM JP-8 PRODUCTION
SUBTASKS 1.2 & 1.3
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
DATE - JAN. 30, 1988

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1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to maximize the production of type JP-8 aviation turbine fuel from Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- . Tar Oil byproduct stream (47620 #/hr, 3182 BPSD) is charged to the hydrotreater (Area 100).
- . The hydrotreater is a 3 stage expanded bed type process which removes 99% + of the sulfur, nitrogen, and oxygen compounds and begins the conversion of 550°F+ material. The hydrotreater adds a large quantity of hydrogen to the feed (4075 SCF/bbl) which results in a high heat of reaction. An expanded bed type reactor was chosen to both control and utilize the heat of reaction. Three stages were used to both control the temperature rise as well as to obtain the high efficiency associated with staging a back-mixed reactor.
- . The hydrotreater produces 6 streams:
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H₂ and CH₄.
 - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.
 - Unstabilized naphtha which is sent to the combined naphtha stabilizer in the hydrocracker (area 200). After stabilization, to control vapor pressure, the naphtha is sent to the main boiler in the SNG plant.
 - JP-8 turbine fuel which is combined with JP-8 produced in the hydrocracker (area 200) and sent to storage.
 - 550°F+ unconverted bottoms product which is sent to the hydrocracker (area 200).
 - Wastewater containing, NH₄OH and NH₄HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H₂S and NH₃.

- Approximately 950 #/day of spent catalyst which is shipped to a catalyst reclaimer in the same drums that the catalyst is received in.
- . The 550°F+ unconverted stream from the expanded bed hydrotreater (area 100) is charged to the fixed bed hydrocracker (area 200). The hydrocracker converts this material to naphtha and JP-8 turbine fuel. For this service a 5 stage unit was chosen with 65% conversion per pass. This unit also includes a naphtha stabilizer which stabilizes both the naphtha produced in the hydrotreater and hydrocracker.
- . The hydrocracker produces 4 streams in addition to JP-8
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit of the SNG plant for recovery of the H₂ and CH₄.
 - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.
 - Stabilized naphtha which is sent to the main boiler in the SNG plant.
 - A small sour water stream which is sent to the PHOSAM unit in the SNG plant or alternatively used as part of the injection water to the hydrotreating plant.
- . Hydrogen make-up for both the Hydrotreater and the Hydrocracker is supplied from a PSA Hydrogen Unit. High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining purge gas is available at a low pressure (5 psig) which has a fuel value of about 565 BTU/ft³. This H₂, CO & CH₄ rich gas is recompressed into the methanation unit of the SNG plant.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. Detailed material balances for each unit can be found in appendixes A&B. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas and naphtha produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

3182 BPSD of Tar Oil
1978 BPSD of #6 Fuel Oil
11.07 MMSCFD equivalent SNG product loss due to the syn gas feed
to the PSA unit.

Products

2490 BPSD of JP-8 turbine fuel
7.37 MMSCFD equivalent SNG product credit due to HDT, HDC & PSA
purge gas reinjection into SNG plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	1978 BPSD
SNG Equivalent	
of Syn Gas & Purge Gas	3.7 MM SCFD
Power	5450 kW
Cooling Water	2040 GPM (30 ⁰ F rise)
Process Water	20.5 GPM

In addition the process exports 6945 #/hr of 100 psig saturated
steam which was credited against boiler requirements.

Figure 1:Case 3-Maximum JP-8 Production

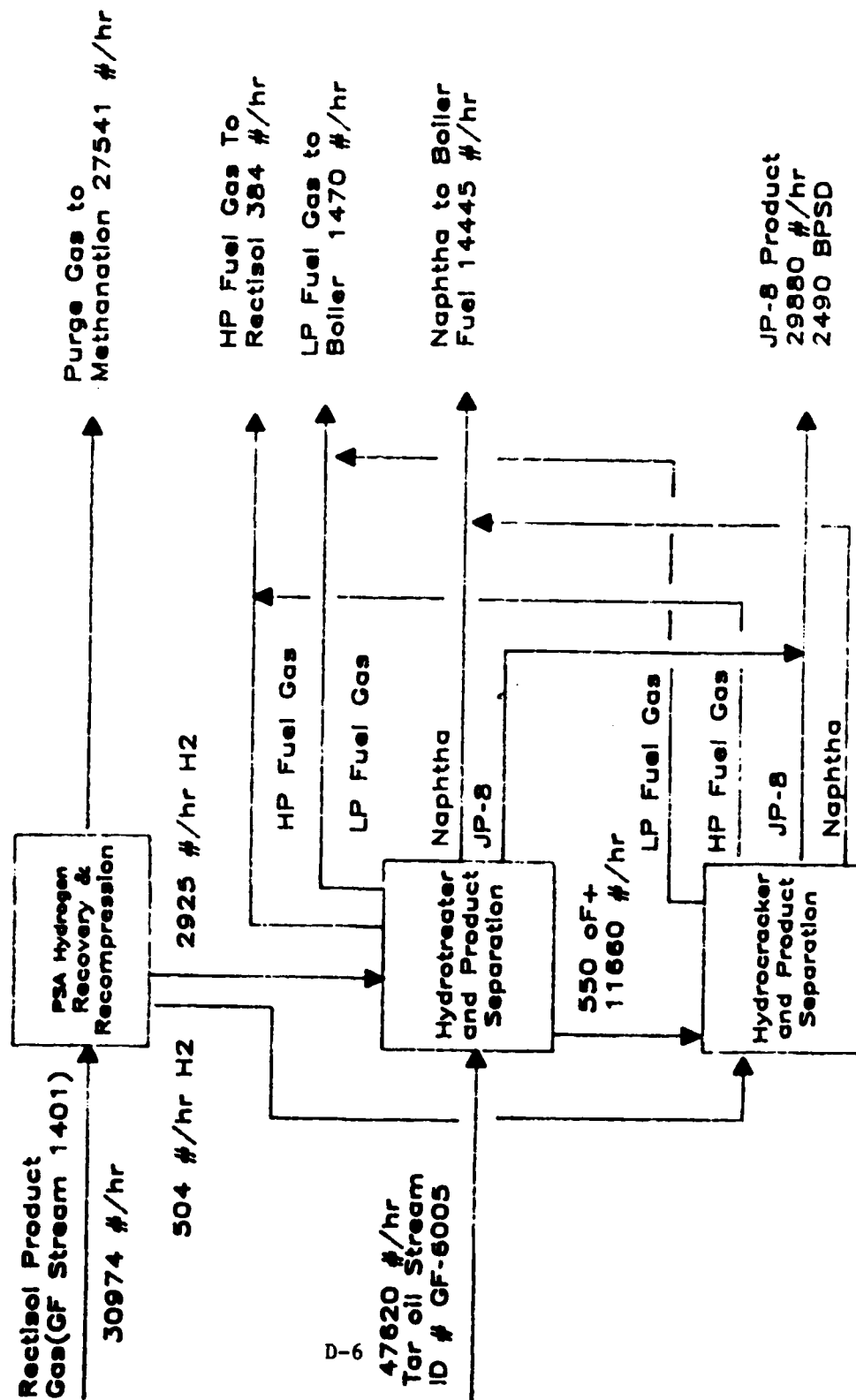


Table 1.: Great Plains Case 3: Maximum JP8 Production

Tar Oil Feed====>	47620	#/hr	3182	BFSD
JP-8 Product====>	29880	#/hr	2490	BFSD
SN6 Product Loss=>	6275	#/hr	3.7	MMSCFD
Fuel Oil Makeup==>	27386	#/hr	1978	BFSD

Expanded Bed Hydrotreater

Comp.	Wt %	Grav	#/hr	#Mole/hr	BFSD
Feeds					
H2	6.14		2925	1450.9	
Oil	100.00	1.0268	47620		3182
Total	106.14		50545		
Products					
Purge Gas	0.10		48	14.9	
Fuel Gas	1.87		891	43.9	
Naphtha	15.17	0.7206	9127		823
JP-8	50.29	0.8720	23950		1975
SSO opF+	24.45	0.9697	11660		825
H2S in SW	8.73		4159	231.1	
H2S in SW	0.43		205	6.0	
NH3 in SW	1.06		505	29.7	
Total	106.14		50545		3623

Fixed Bed Hydrocracker

Comp.	Wt %	Grav	#/hr	#Mole/hr	BFSD
Feeds					
H2	4.32		504	250.0	
SSO opF+	100.00	0.9697	11660		825
Total	104.32		12164		
Products					
Purge Gas	2.88		336	113.9	
Fuel Gas	2.12		247	13.7	
Naphtha	48.46	0.7148	5650		542
JP-8	50.86	0.7900	5930		515
H2S in SW	0.003		0.4	0.01	
NH3 in SW	0.003		0.4	0.02	
Total	104.32		12164		1057

Naphtha Stabilizer

Comp.	Wt %	Grav	#/hr	#Mole/hr	BFSD
HDT Nap	61.76	0.7606	9127		823
HCR Nap	38.24	0.7148	5650		542
Stab Nap	97.75	0.7346	14445		1349
Fuel Gas	2.25		332	7.8	

PSA Hydrogen Recovery Unit (86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	6.63	5.55	901.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	6.63	5.55	805.15
#Mol/hr							
Feed Gas	1977.6	582.7	46.3	507.5	9.9	6.0	3130.2
Prod. H2	1700.9	0.2	0.0	0.0	0.0	0.0	1701.1
Purge Gas	276.9	582.6	46.3	507.5	9.9	6.0	1429.1
#/hr							
Feed Gas	3987	16320	2038	8143	296	190	20974
Prod. H2	3429	4	0	0	0	0	3433
Purge Gas	558	16316	2038	8143	296	190	21511

Fuel Gas Generated in Hydrotreating and Hydrocracking

Component	#/hr	#Mol/hr	MMBTU/hr
HDTR FG Produced	891	43.9	16.0
HCR FG Produced	247	13.7	4.4
Stab FG Produced	332	7.8	6.0
Total Fuel Gas	1470	65.4	26.5

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	558	276.9	324	34.0
CO	16316	582.6	321	70.9
CO2	2038	46.3	0	0.0
C1	8143	507.5	1010	194.3
C2	296	9.9	1769	6.6
N2+Ar	190	6.0	0	0.0
Total	27541	1429.1	565	305.8

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft ³	BPSD
Tar Oil	-47620	17000	-805.5			-3182
Fuel Gas	1470	18000	26.5	0.6	1068	
Naphtha	14445	20040	289.5			1349
Export Steam	6945	1000	6.9			
Fuel Oil to Boiler	27036	18000	486.7			1950
Total	2277		0.0	0.6		120
Fuel Oil to Process Heaters	350	18000	6.3			25

Net Changes in SNG Production	EDV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	11.02	75124
SNG Credit for PSA Purge gas	6.90	34292
SNG Credit for H ₂ Purge gas	0.47	3091
Total SNG Production Loss	3.65	

Water Balance Hydroheater

Component	#/hr					Total
	H ₂ O	H ₂ S	NH ₃	NH ₄ HS	NH ₄ OH	
Reaction Gases	4159	205	505			4869
Reaction Solution	3733			308	829	4869
Stripping Steam	1225					1225
Softened Water	9281					9281
HDT Sour Water	14239			308	829	15075

Water Balance Hydrocracker

Component	#/hr					Total
	H ₂ O	H ₂ S	NH ₃	NH ₄ HS	NH ₄ OH	
Reaction Gases		0.4	0.4			0.7
Reaction Solution				0.5	0.4	0.9
Stripping Steam	225.1					225.1
HCR Sour Water	225.1			0.5	0.4	226.0

Total Sour Water

Component	#/hr					Total
	H ₂ O	H ₂ S	NH ₃	NH ₄ HS	NH ₄ OH	
Total Sour Water	14464			308	829	15601

2.0 PROCESS DESCRIPTION

2.1 Hydrotreater (Area 100)

Operating conditions for the hydrotreater were provided to Lummus by Amoco and these conditions are presented in Table 2.1. The basic processing step selected was the expanded bed hydrotreater (LC Fining) system. Due to the extremely high exothermic heat of reaction it was necessary to use 3 reactors in with interstage cooling. Referring to drawing D5571-30101 and the material balance printouts (Appendix A) the flow is as follows:

- . Feed Tar Oil is charged into the hydrotreater from day tank FA-101 and through charge pump GA-101 and preheater exchanger EA-101.
- . The preheated charge oil is combined with feed hydrogen gas from heater BA-101. Preheat of the oil is limited to 550°F to prevent cracking. The preheated mixture is then charged to the first reactor DC-101A.
- . The expanded bed reactor DC-101A approaches isothermal conditions in which the heat of reaction is used to heat the feed up to 700°F.
- . The effluent from DC101A is cooled in exchanger EA-101 and combined with recirculating hydrogen from recycle hydrogen gas compressor GB-102. The combined mixture is charged into the second reactor where the heat of reaction increases the temperatures to 700°F.
- . The effluent from DC101B is cooled in exchanger EA-104 and combined with recirculating hydrogen gas from recirculating compressor GB-102. This mixture is charged into the third reactor DC-101C.
- . The effluent from DC101C goes to the high temperature/high pressure separator FA-103. Hot liquid from FA-103 flows to the hydrotreater fractionation DA-101. The vapors from FA-103 flows through exchangers EA-102 and EA-105 and then through air cooler EC-101. Process water is injected prior to EC-101 to convert the H₂S and NH₃ in the gas to an aqueous NH₄OH/NH₄HS solution.
- . Exchangers EA-104 and EA-105 are part of a circulating hot oil belt which allows for the generation of steam from waste heat in the high pressure loop without having the problem of a hydrocarbon leak from the high pressure system into the steam system.

2.1 Hydrotreater - Cont'd

- . The cooled gas then passes into the High Pressure/Low Temperature Separator FA-104 where hydrogen rich gas is taken as an overhead product. A purge stream of this high pressure gas is taken (to purge H₂ and light gases from the loop) and sent to the Rectisol Unit 1400 in the SNG plant to recover the hydrogen in the purge gas. The remaining gas is recirculated to reactors DC-101B and DC-101C.
- . The water phase from separator FA-104 goes to the PHOSAM Unit in the SNG plant to recover the H₂S and NH₃.
- . The hydrocarbon phase from separator FA-104 is preheated in exchanger EA-105 and is combined with the hot liquid from FA-103 and charged to the HDT Fractionator DA-101. Fractionator DA-101 produces 550°F+ product (which is sent to hydrocracking, area 200), JP-8 (which is sent to storage), and unstabilized naphtha (which is sent to the naphtha stabilizer in the hydrocracking area 200).
- . Catalyst is replaced every third day in each reactor so that one reactor is receiving and withdrawing catalyst each day. Catalyst is added and replaced by the catalyst handling system.
- . Waste heat is converted to 100 psig saturated steam in exchangers EA-107 and EA-108. This steam is used for stripping in DA-101 and in the hydrocracking area 200 for stripping steam. There is an excess of about 6945 #/hr which is exported to the SNG steam system.

Table 2.1 Hydrotreater Conditions

Case 3 Maximum JP-8 Operation

Reactor Type	Expanded Bed
Number of Reactors	3
Reactor Temperature	700°F
Temp. rise/stage	225°F max.
Ratio of H ₂ in Feed to Chemical H ₂	2.0 min.
Catalyst Replacement	0.30 #/Bbl

2.2 Hydrocracker (Area 200)

Operating conditions for the hydrocracker were provided to Lummus by AMOCO and these conditions are presented in Table 2.2. The basic processing step selected was a 5 bed fixed bed reactor system with recycle of unconverted 550°F+ material. Beds 1 and 2 use a catalyst that is most active for sulfur, nitrogen and oxygen removal while beds 3,4,5 use a catalyst that is most active for hydrocracking. Referring to drawing D5571-30201 and the material balance printouts (Appendix B) the flow is as follows:

- Hydrotreated 550°F+ material from the hydrotreater (Area 100) enters the system from day tank FA-201 through feed pump GA-201 and is preheated in exchanger EA-201. The preheated feed is combined with unconverted bottoms from fractionator DA-201 (approximately 35% of the feed is recycled). The combined oils are then mixed with hot hydrogen coming from heater BA-201 and charged to the reactor.

The combined feed to the first bed in the reactor is 662°F.

- In the first bed the temperature rises to about 675°F. Quench hydrogen is added to cool the effluent from the first bed to about 648°F. In each of the remaining beds quench hydrogen is added to cool the beds. The inlet and outlet temperatures from each bed are as follows:

	Inlet	Outlet
Bed 1	662	675
Bed 2	648	676
Bed 3	652	677
Bed 4	656	680
Bed 5	661	684

- The reactor effluent passes into the high temperature/high pressure separator FA-202. Vapors from FA-202 are cooled in EA-201, EA-202 and then air condenser EC-201. Water is injected into the condenser EC-201 to dissolve H₂S and NH₃ into a NH₄OH/NH₄HS solution. This solution is sent to the PHOSAM unit in the SNG plant.
- The cooled vapors pass into separator FA-203 and the overhead hydrogen rich gas is divided with the major portion being used as recycle gas to the reactors via compressor GB-202 and heater BA-201. A small portion of the gas is purged from the system as high pressure purge gas which goes to the Rectisol unit in the SNG plant.

2.2 Hydrocracker (Area 200)

- . The hydrocarbon phase from separator FA-203 is heated in exchanger EA-202, combined with the hot oil from separator FA-202 and charged to fractionator DA-201.
- . Fractionator DA-201 produces unstabilized naphtha (which is charged to naphtha stabilizer DA-203), JP-8 (which is sent to product storage after cooling), 550°F+ oil (which is recycled to reactor DC-201) and fuel gas (which flows to the boiler in the SNG plant).
- . Unstabilized naphtha from DA-201 is combined with unstabilized naphtha from area 100 and charged to naphtha stabilizer DA-203 after being preheated in exchanger EA-205. Heat for reboiling the naphtha stabilizer is obtained by heat exchange with the hot jet fuel product.
- . Fuel gas from the naphtha stabilizer is combined with fuel gas from FA-206 and is routed to the SNG boilers.
- . The stabilized naphtha is cooled and sent to storage. It is also sent to the SNG boilers to be used as fuel.

Table 2.2 Hydrocracker Conditions

Reactor Conditions	5 Fixed Beds
Catalyst, % of Total & Type	
Bed 1	10%, HDS/HDN/HC
Bed 2	22.5%, HDS/HDN/HC
Beds 3-5	22.5% HC
WHSV, hr ⁻¹	1.5
Average Reactor Temp.	675°F
Temperature Increase	
Bed 1	15°F
Bed 2-5	40°F
Heat of Reaction	20,000 BTU/#Mole H ₂
Reactor Pressure	
Inlet	1200 psig
Outlet	1175 psig
Recycle Rate H ₂	13,000 scf/Bbl
Conversion/Pass	50%
Catalyst Replacement	3 years @ \$6/#

2.3 PSA Hydrogen Unit & Recompression (Area 300)

- 2.3.1 Hydrogen for both the hydrotreater and the hydrocracker will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig
Temp.	65 °F
Composition	mol%

H2	63.19
CO	18.61
CO2	1.48
CH4	16.21
C2H6	0.31
COS, H2S, CS2	< 0.01
N2 + Ar	0.19
H2O	< 0.01

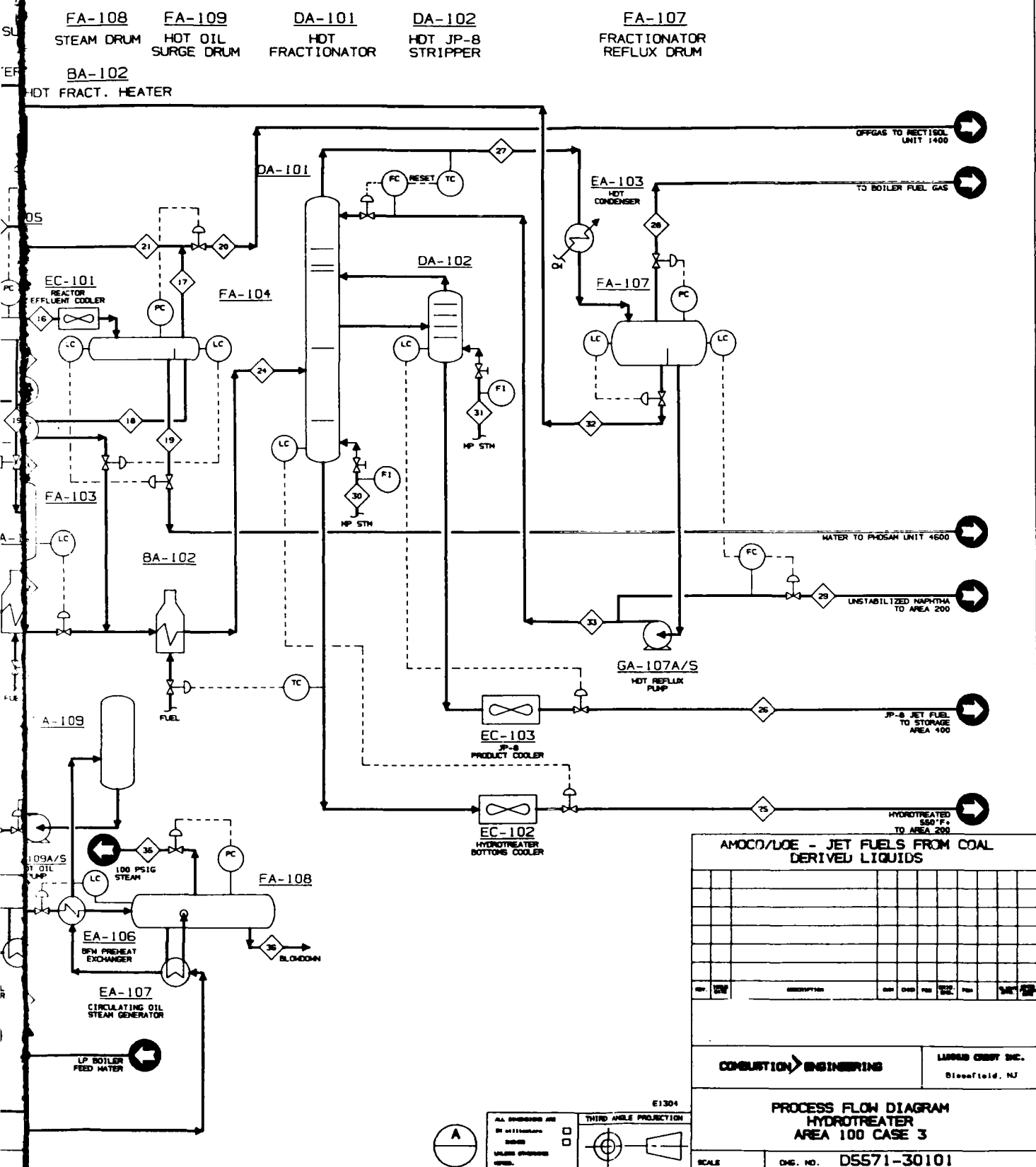
The PSA unit selectively absorbs all components except H2 and produces a 99.99% vol. purity stream at about 345 psig and 80°F. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100°F
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

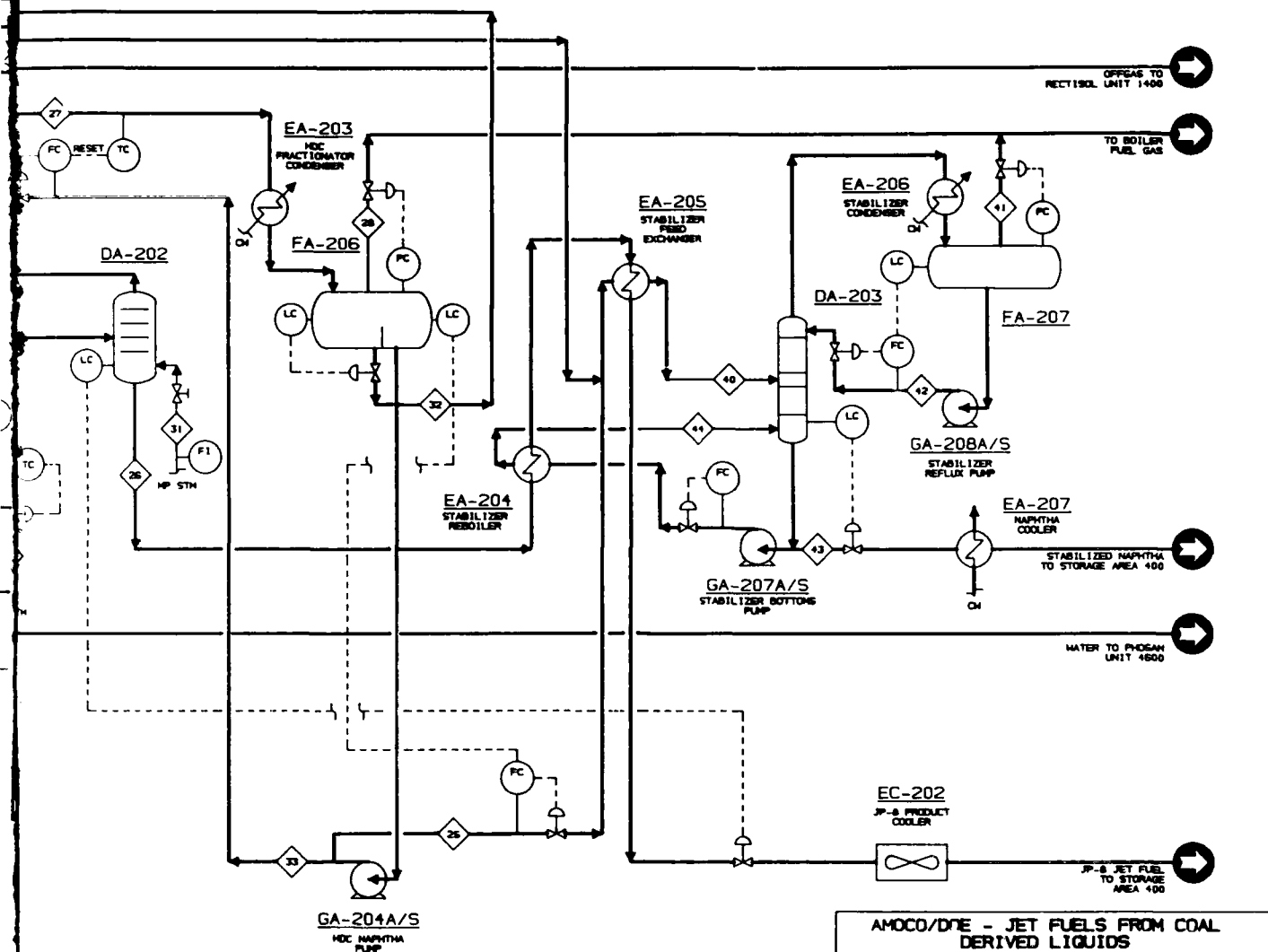
At the conditions given a 10 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

The system uses 10 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continuously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-30301 presents a schematic of a Union Carbide Polybed PSA unit.

- 2.3.2 The purge gas is recompressed to 375 psia and sent to the methanation unit of the SNG plant.



FA-207
STABILIZER
REFLX DRUM



AMOCO/DNE - JET FUELS FROM COAL
DERIVED LIQUIDS

[illegible]

LAGOS CREDIT INC.
Bloomfield, NJ

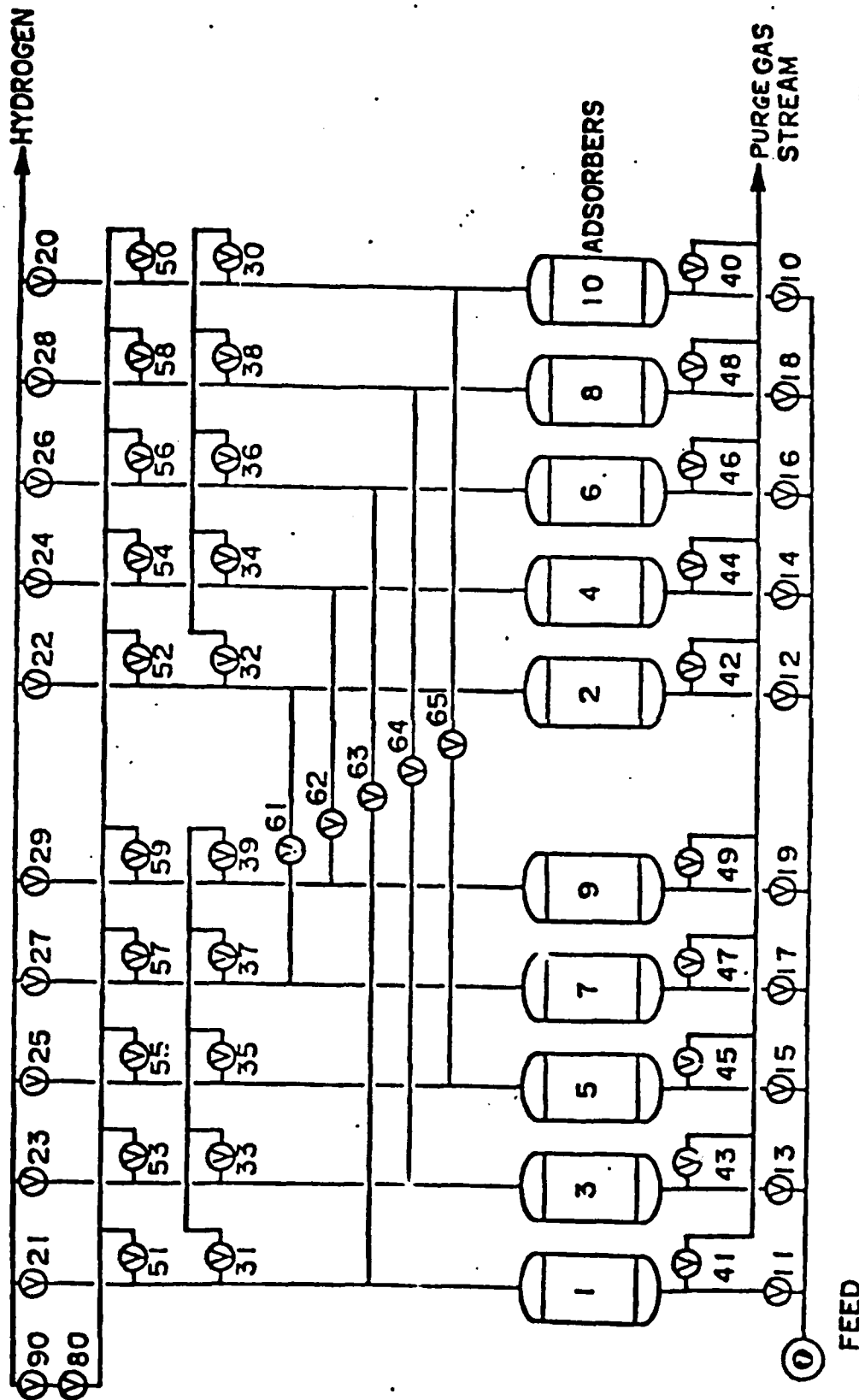
SCALE	DWG. NO. D5571-30201



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SCHEMATIC FLOW SHEET - POLYBED PSA UNIT



THE LUMBER COMPANY	THE LUMBER COMPANY
PSA HYDROGEN UNIT	PSA HYDROGEN UNIT
CURRENT AMOCO/DOE	CURRENT AMOCO/DOE
PROJ NO 5571	PROJ NO 5571
CASE 3	CASE 3

TYPICAL ARRANGEMENT
NUMBER OF ADSORBERS FOR THIS CASE = 10

Δ	FOR SUBTASK 1.2	ML	ES	PROD	REGD	DATE	TIME
Δ	FOR SUBTASK 1.2	ML	ES	PROD	REGD	DATE	TIME

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.1 Equipment List

CASE 3 - MAXIMUM JP-8

<u>AREA 100</u>	-	<u>HYDROTREATER</u>
<u>TAG NO.</u>	-	<u>DESCRIPTION</u>
BA-101		HDT Makeup Gas Heater
DA-101		HDT Fractionator
DA-102		JP-8 Stripper
DC-101A,B,C		Hydrotreater Reactors
EA-101		HDT Reactor Feed/Effl. Exch.
EA-102		Fract. Feed Exch.
EA-103		HDT Condenser
EA-104		HDT Reactor Int. Stg. Clr.
EA-105		Waste Heat Exchanger
EA-106		BFW Preheat Exch.
EA-107		Circulating Oil Stm. Gen.
EC-101		Reactor Effl. Cooler
EC-102		Hydrotreater Btms. Cooler
EC-103		JP-8 Product Cooler
FA-101		Tar Oil Surge Drum
FA-102		HDT Makeup Gas KO Drum
FA-103		HT/HP Separator
FA-104		LT/HP Separator
FA-105		Recycle Gas KO Drum
FA-107		Fractionator Reflux Drum
FA-108		Steam Drum
FA-109		Hot Oil Surge Drum
GA-101A/S		Tar Oil Feed Charge Pump
GA-103A/S		Wash Water Pump
GA-107A/S		HDT Reflux Pump
GA-109A/S		Hot Oil Pump
GA-110A/B/C		Reactor Recycle Pump
GB-101A/B		H ₂ Makeup Compr.
GB-102A/B		Recycle Gas Compr.

CASE 3 - MAXIMUM JP-8 - Cont'd

<u>AREA 200</u>	<u>HYDROCRACKER</u>
<u>TAG. NO.</u>	<u>DESCRIPTION</u>
BA-201	HDC Reactor Gas Heater
DA-201	HDC Fractionator
DA-202	HDC JP-8 Stripper
DA-203	Naphtha Stabilizer
DC-201	HDC Reactor
EA-201	HDC Reactor Feed/Effl. Exch.
EA-202	HDC Fract. Feed Exch.
EA-203	HDC Fract. Condenser
EA-204	Stabilizer Reboiler
EA-205	Stabilizer Feed Exch.
EA-206	Stabilizer Condenser
EA-207	Naphtha Cooler
EC-201	Reactor Effl. Cooler
EC-202	JP-8 Product Cooler
FA-201	HDC Feed Surge Drum
FA-202	HT/HP Separator
FA-203	LT/HP Separator
FA-204	Recycle Gas KO Drum
FA-205	Makeup Gas KO Drum
FA-206	HDC Fract. Reflux Drum
FA-207	Stabilizer Reflux Drum
FA-208	Fuel Oil Day Tank
GA-201A/S	HDC Feed Pump
GA-202A/S	Wash Water Pump
GA-203A/S	HDC Recycle Pump
GA-204A/S	HDC Naphtha Pump
GA-207A/S	Stabilizer Btms Pump
GA-208A/S	Stabilizer Reflux Pump
GA-209A/S	Fuel Oil Pump
GB-201A/B	Makeup Gas Compr.
GB-202A/B	Recycle Gas Compr.
<u>AREA 300</u>	<u>PSA HYDROGEN UNIT & RECOMPRESSION</u>
FA-301	Purge Gas Surge Drum
GB-301	Purge Gas Compressor
PA-301	PSA Hydrogen Unit Package

CASE 3 - MAXIMUM JP-8 - Cont'd

<u>TAG NO.</u>	<u>DESCRIPTION</u>
<u>AREA 400</u> -	<u>STORAGE AREA</u>
FB-401	Jet Fuel Storage Tank
FB-402	Naphtha Storage Tank
FB-403	Fuel Oil Storage Tank
GA-401A/S	Tar/Tar Oil Feed Pump
GA-403A/S	Fuel Oil Transfer Pump
GA-404A/S	Naphtha Transfer Pump
<u>AREA 500</u> -	<u>CATALYST HANDLING</u>
<u>TAG NO.</u>	<u>DESCRIPTION</u>
FA-501	Catalyst Oil Drum
FA-502	Catalyst Storage Hopper
FA-503	Catalyst Transfer Vessels
FA-504	Spent Catalyst Vessel
FL-501	Catalyst Screen
GA-501A/S	Catalyst Transfer Pump
GA-502A/S	Catalyst Oil Pump

3.2 Cost Estimate

3.2.1 Basis of Estimate

The estimate is an equipment factored type estimate using the equipment sizes & specifications developed for this project. The equipment unit pricing is based on return data for high pressure equipment purchased for various hydrotreater/hydrocracker projects. The unit pricing is somewhat conservative compared to world wide markets of 2-3 years ago, however, the exchange rate decline during this period will lead to higher purchase prices.

The commodity materials & subcontracts are ratioed from the equipment costs using factors considering the high pressure processing, the size of the units, and the location of the plant.

The labor and indirects also are factored considering process, sizing, and location.

Engineering costs are based on the equipment count times the historical number of manhours per equipment item, and the current average engineering selling rate.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs.

Excluded from this estimate are:

- Spare Parts
- Start-Up
- Insurances & Taxes
- Permits
- Royalties on Processing Technology Knowhow

3.2.2 Estimate Summary

(Thousands of \$)

Case 3

Area 100 Hydrotreater	\$20,702
Area 200 Hydrocracker	10,012
Area 300 PSA & Recompression	8,300
Area 400 OSBL	4,500
Area 500 Catalyst Handling	1,285
Total	<u>\$44,799</u>

3.2.3 Estimate Breakdown (Area 100) All Values in Thousands

	<u>Equipment</u>	<u>\$ Value</u>	<u>%, Comm</u>	<u>\$ Comm.</u>
<u>Items</u>	<u>Type</u>			
1	Heaters	50	120	60
2	Towers	39	140	55
-	Internals	7	-	-
3	Reactors	1650	70	1155
13	Exchangers	582	70	407
3	Air Coolers	113	100	113
8	Vessels	388	85	330
11	Pumps	739	80	591
4	Compressors	1700	60	1020
	Special	-		
45	Total	\$5268		\$3731
	Equipment		5268	
	Commodities		3731	
	Labor		2765	10% Equip. 60% Comm.
	Indirects		2765	100%
	Office		2723	45 pcs x 1100 x \$55-
	Subtotal		17,252	
	Contingency		3,450	20%
	Total		\$20,702	

3.2.3 Estimate Breakdown - Cont'd

Area 200

	<u>Equipment</u>	<u>\$ Valve</u>	<u>% Comm</u>	<u>\$ Comm.</u>
<u>Items</u>	<u>Type</u>			
1	Heaters	130	120	156
3	Towers	54	140	70
-	Internals	9	-	-
1	Reactors	325	85	276
8	Exchangers	188	100	188
2	Air Coolers	46	110	50
8	Vessels	258	100	258
14	Pumps	152	120	182
4	Compressors	550	100	550
	Special	-		
41	Total	\$1712		\$1730

Equipment 1712

Commodities 1730

Labor 1210 10% Equip. 60% Comm.

Indirects 1210 100%

Office Subtotal 2481 41 pcs x 1100 x \$55-
8343

Contingency 1669 20%
Total \$10,012

Area 300

PSA-unit 20 mm SCFD budget quote \$3500

PSA unit 14 mm SCFD 2500

Installation 50% 1300

Subtotal \$3800

Compressor 3400 1700 x 2.0 T.I.C.
Drum 300 100 x 3.0 T.I.C.
Subtotal \$7500

Contingency 800 10%

Total \$8300

3.2.3 Estimate Breakdown - Cont'd

Area 400

Case 1 TIC = \$5100

For Case 3 Deduct Equipment 3 x 150 = 450
Piping 200
Total Deduct \$650

Case 3 Total - \$4500

Area 500

	<u>Equipment</u>	<u>\$ Value</u>	<u>% Comm</u>	<u>\$ Comm.</u>
<u>Items</u>	<u>Type</u>			
4	Vessels	105	120	126
4	Pumps	48	120	58
8	Total	\$153		\$184
	Equipment		153	
	Commodities		184	
	Labor		125 10% Equip. 60% Comm.	
	Indirects		125 100%	
	Office		484 8 pcs x 1100 x \$55-	
	Subtotal		\$1071	
	Contingency		214 20%	
			\$1285	

4.0 OPERATING COSTS

4.1 Operating Labor

It is estimated that it will require 7 men/shift to operate the plant broken down as follows:

Foreman	1	
Control Room	1	
HDT Operator	2	
HCR Operator	2	
PSA & relief man	1	
	<u>7</u>	Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 5 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	7 positions x 4 people/position	- 28
Supervisor & Admin.		5
QC Technician		1
Maintenance		5
Other (Stores or Janitorial)		1
Total		<u>40</u>

4.2 Utilities

The following utilities have been estimated from the computer simulations:

<u>Utility</u>	<u>Consumption</u>	<u>Cost</u>	<u>\$/SD</u>
#6 Fuel Oil	1978 BPSD	\$16/Bbl (a)	31648
SNG equivalent	3.7 MMSCFD	\$3.80/MM Btu (b)	13780
of Syn Gas & Purge Gas			
Cooling Water	2040 GPM	\$0.155/MGal (c)	456
Power	5450 kW	\$0.04/kWh (c)	5232
Process Water	20.5 GPM	\$0.45/MGal (c)	13

(a) Cut of 1% sulfur \$6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

(b) Memo from D. Daley of Burns & Roe to L. Lorenzo of DOE dated Oct. 20, 1987, reference DPD-87-863.

(c) ANG utility cost information dated 5/87.

4.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

<u>Catalyst</u>	<u>Use</u>	<u>Cost</u>	<u>\$/SD</u>
HDT Cat.	0.30 #/Bbl	\$3.00/#	2864
HCR Cat.	0.0095 #/Bbl	\$6.00/#	47
Inhibitors	50 PPM	\$10/Gal	52
			<u>2963</u>

4.4 Maintenance Supplies

Maintenance supplies for hydrotreating operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units. On this basis the maintenance supplies would be

$$0.00005 \times 44,799,000 = \$2240./SD$$

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the production of JP-8 are proposed to be located to the east of the Rectisol Unit and Main Control Room of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 300' x 200' will be surrounded by an access road and will be divided by a central east-west road. Areas 100 & 500 will be located to the north and Areas 200 & 300 to the south.

A diked storage tank area approx. 360' x 265' will be required for product and fuel oil storage and is proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

5.2 Unit Tie-Ins

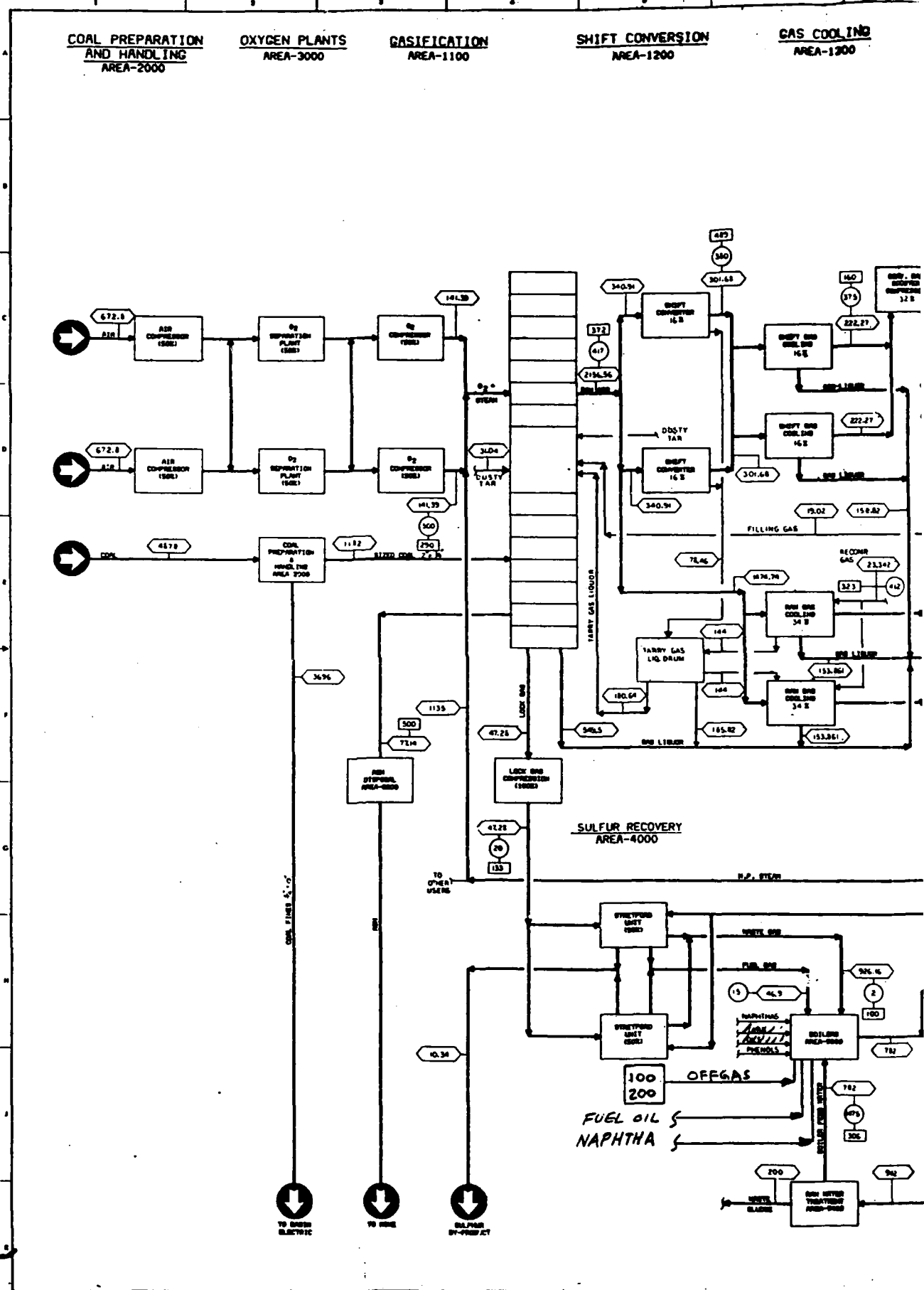
Approximately 2000 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

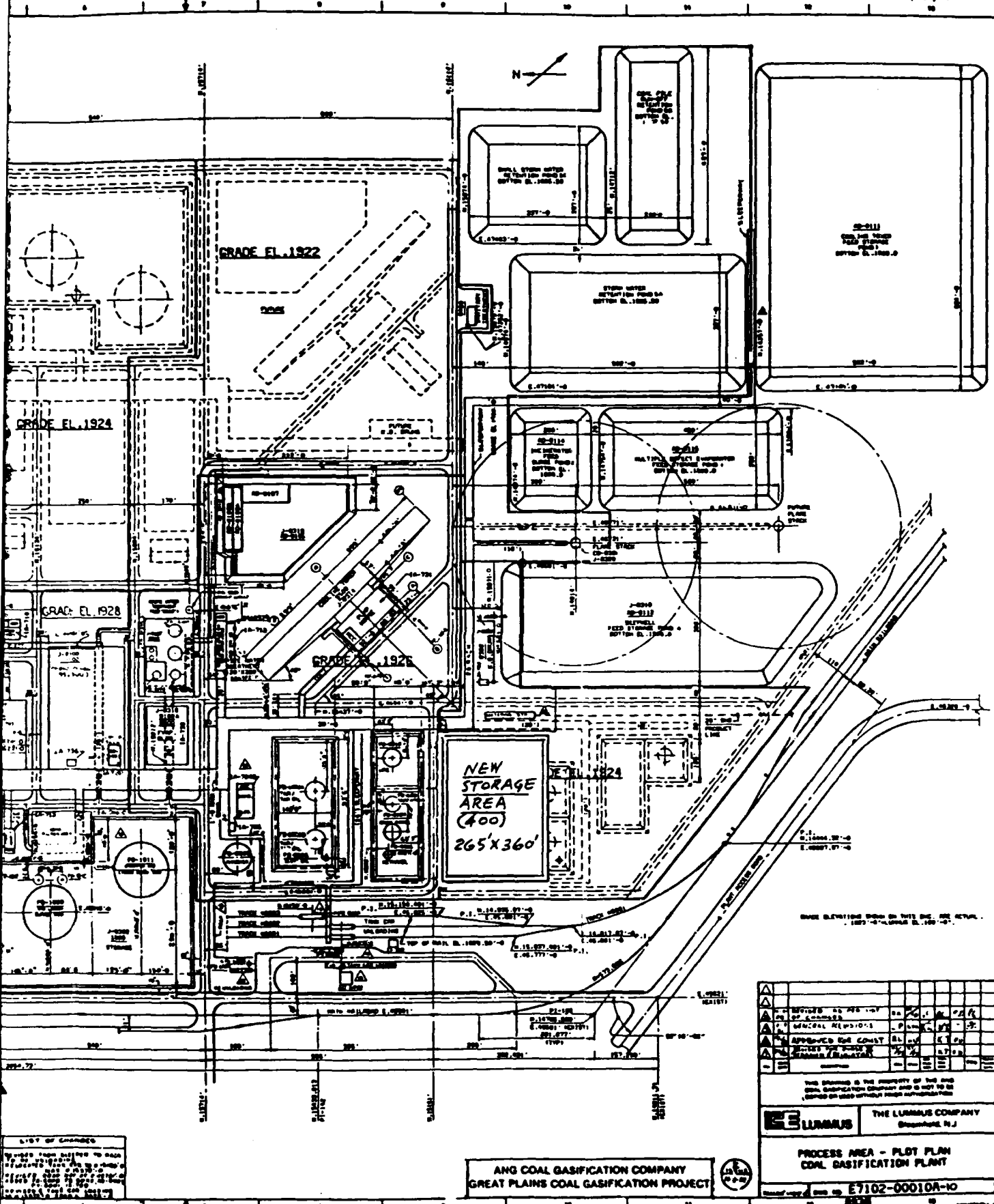
New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

A summary of the lines is shown in table 5.1.

TABLE 5.1
INTERCONNECTING PIPING

<u>EST. SIZE</u>	<u>SERVICE</u>	<u>TO/FROM</u>
4"	Tar/Tar Oil (Elec. Tr.)	Storage
3"	JP-8 Product	Storage
2"	Naphtha Product	Storage
16"	Wet Flare (Trace)	Flare
8"	Synthesis Gas	PSA/Rectisol
6"	Purge Gas	Methanation/PSA
1 1/2"	Off Gas	Rectisol/HDT,HDC
2"	Nitrogen	Main Rack
2"	Plant Air	"
2"	Instr. Air	"
2"	Raw Water (Elec. Tr.)	"
6"	M.P. Steam	"
1 1/2"	Stm Cond.	"
1 1/2"	BFW	"
1 1/2"	Boiler B.D.	"
12"	C. W. Supply & Return	"
2"	Waste Water	Phosam/HDT,HDC
4"	Fuel Oil	Exist TKS/New TKS.
15"	Storm Sewer (9' deep)	Storm Basin
15"	Oily Water Sewer (9' deep)	8100/Process Unit
6"	Sanitary Sewer (9' deep)	8400/Process Unit
10"	Fire Water	Ring Headers





CASE 3

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TYPE A-1
CASE 3
AREA 100

PAGE 13
ML
NOV 17 1997

VERSION 2.01
SIMULATION SCIENCES INC.
PROJECT GP JET FUELS
PROBLEM C3U100

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	1	2	3	4	23	24
STREAM NAME	FEED TAR LIQUID	MAKEUP H2 VAPOR	HOT TAR OIL LIQUID	NET MAKEUP VAPOR	101B REC GAS VAPOR	101C REC GAS VAPOR
STREAM PHASE	LIQUID	VAPOR	LIQUID	VAPOR	VAPOR	VAPOR
TEMPERATURE, DEG F	154.0000	70.0000	550.0000	395.9943	151.1667	151.1667
RATE LB MCL/HR	304.8516	1450.0000	304.0535	1449.9998	728.4996	672.4612
RATE LB /HR	47707.5703	2925.2324	47707.5703	2925.2324	2488.5923	2297.1611
ENTHALPY MM BTU /HR	2.0518	-3.0145	11.3451	0.1755	-0.9060	-0.8363
ENTHALPY BTU /LB	43.0074	-1030.5194	237.3040	59.9865	-364.0686	-364.0686
MOLECULAR WEIGHT....	156.4934	2.0174	156.4934	2.0174	3.4161	3.4160
*** VAPOR PHASE ***						
RATE LB /HR	0.0000	2925.2324	0.0000	2925.2324	2488.5923	2297.1611
STD. RATE MM FT3/DAY	0.00	13.21	0.00	13.21	6.63	6.12
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4460	2.1428	2.1428
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	3.4161	3.4160

*** LIQUID PHASE ***						
RATE LB /HR	47707.5703	0.0000	47707.5703	0.0000	0.0000	0.0000
ACT. RATE BBL/DAY	3269.42	0.00	3824.95	0.00	0.00	0.00
STD. LV RATE BBL/HR	133.10	0.00	133.10	0.00	0.00	0.00
CP, BTU /LB F	0.4049	0.0000	0.5913	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	156.4934	0.0000	156.4934	0.0000	0.0000	0.0000
ACT. DENS LB /FT3	62.3749	0.0000	57.3155	0.0000	0.0000	0.0000
STD. API GRAVITY....	6.5566	0.0000	6.5896	0.0000	0.0000	0.0000
*** DRY BASIS ***						
RATE LB /HR	46753.7656	0.0000	46753.7656	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	185.5983	0.0000	185.5983	0.0000	0.0000	0.0000
UOP K	9.9510	0.0000	9.9530	0.0000	0.0000	0.0000
FLASH POINT, DEG F	142.3663	0.0000	142.3663	0.0000	0.0000	0.0000
CRIT. TEMP, F	965.2936	0.0000	965.2936	0.0000	0.0000	0.0000
CRIT. PRES, PSIA	424.8965	0.0000	424.8965	0.0000	0.0000	0.0000

*** VAPOR PHASE ***						
RATE LB /HR	0.0000	2925.2324	0.0000	2925.2324	2477.1152	2286.5669
STD. RATE MM FT3/HR	0.00	0.55	0.00	0.55	0.28	0.25
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4460	2.1505	2.1506
MOLECULAR WEIGHT....	0.0000	2.0174	0.0000	2.0174	3.4033	3.4033

VISCOSITY, CP						
	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /HR	46753.7656	0.0000	46753.7656	0.0000	0.0000	0.0000
ACT. RATE BBL/DAY	3262.71	0.00	3736.27	0.00	0.00	0.00
STD. LV RATE BBL/HR	0.3949	0.0000	0.5790	0.0000	0.00	0.00
CP, BTU /LB F	185.5983	0.0000	185.5983	0.0000	0.0000	0.0000
MOLECULAR WEIGHT....	185.5983	0.0000	185.5983	0.0000	0.0000	0.0000
ACT. DENS LB /FT3	62.4010	0.0000	57.4497	0.0000	0.0000	0.0000
STD. API GRAVITY....	6.5177	0.0000	6.5177	0.0000	0.0000	0.0000

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	10	11	12	13
STREAM NAME	101A INLET	101A OUTLET	101B OUTLET	101C OUTLET
STREAM PHASE	FIXED	MIXED	MIXED	MIXED
TEMPERATURE, DEG F	467.2208	700.0000	700.0000	700.0000
RATE LB MOL/HR	1754.8521	1363.4368	1714.6738	2037.5083
RATE LB /HR	50632.9047	50632.9047	53122.7031	55423.3672
ENTHALPY MM BTU /HR	11.5204	22.5462	20.3515	29.9376
ENTHALPY BTU /LB	227.5274	445.2932	496.0495	540.1617
MOLECULAR WEIGHT.....	28.8530	37.1357	30.9812	27.2016
*** VAPOR PHASE ***				
RATE LB /HR	5532.7441	18865.8438	32405.3750	47019.6406
STD. RATE MM FT3/DAY	13.41	10.44	14.31	18.02
CP, BTU /LB F	1.9962	0.8624	0.8264	0.8150
MOLECULAR WEIGHT.....	3.7569	10.4446	20.6213	23.7685
*** LIQUID PHASE ***				
RATE LB /HR	45100.0459	31767.6367	20717.3203	8403.7207
ACT. RATE BBL/DAY	3645.26	3221.31	2161.44	901.53
STD. LV RATE BBL/HR	129.51	55.52	54.67	27.26
CP, BTU /LB F	0.5512	0.6658	0.5971	0.7214
MOLECULAR WEIGHT.....	159.8336	145.9953	146.6516	141.7721
ACT. DENS LB /FT3	52.0860	42.1546	40.9716	39.8324
STD. API GRAVITY.....	9.5333	17.3274	23.0012	29.0798
*** DRY BASIS ***				
RATE LB /HR	44961.9359	31526.7969	20533.1992	8322.6650
MOLECULAR WEIGHT.....	163.7950	154.3732	154.3628	151.9371
UOP K	10.0351	10.4339	10.7379	11.0075
FLASH POINT, DEG F	10.7020	-1.0073	-19.3667	-32.2335
CRIT. TEMP, F	788.6436	660.3578	623.0140	579.9640
CRIT. PRES, PSIA	366.2634	349.6859	359.6001	329.3538
*** VAPOR PHASE ***				
RATE LB /HR	4717.1543	16974.6434	29310.3320	42758.4375
STD. RATE MM FT3/DAY	0.54	0.39	0.53	0.66
CP, BTU /LB F	2.2542	0.4020	0.6007	0.8457
MOLECULAR WEIGHT.....	3.3047	10.3032	20.9412	24.5499
*** LIQUID PHASE ***				
RATE LB /HR	8294.7793	8294.7793	8294.7793	8294.7793
ACT. RATE BBL/DAY	684.95	684.95	684.95	684.95
CP, BTU /LB F	0.0000	0.0000	0.0000	0.0000
MOLECULAR WEIGHT.....	0.0000	0.0000	0.0000	0.0000
ACT. DENS LB /FT3	0.0000	0.0000	0.0000	0.0000
STD. API GRAVITY.....	0.0000	0.0000	0.0000	0.0000

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STREAM ID	15	16	17	19	20
STREAM NAME	W2C WASH EFF + W2D	COLD SEP VAP	COLD SEP LIQ	SOUR H2O PURGE GAS	
STREAM PHASE		VAPOR	LIQUID	LIQUID	VAPOR
TEMPERATURE, DEG F		120.0000	120.0000	120.0000	
RATE LB MOLS/HR		1416.9827	303.4501	16.0176	
RATE LB /HR		4840.5156	37544.5781	54.7176	
ENTHALPY MM BTU /HR		-2.0745	0.4793	-0.0235	
ENTHALPY BTU /LB		-423.5753	12.7454	-428.5779	
MOLECULAR WEIGHT....		3.4161	123.7257	3.4161	
*** VAPOR PHASE ***					
RATE LB /HR		4840.5156	0.0000	54.7176	
STD. RATE MM F13/DAY		12.97	U.UUU	0.15	
CP, BTU /LB F		2.1473	0.0000	2.1464	
MOLECULAR WEIGHT....		3.4161	0.0000	3.4161	
*** LIQUID PHASE ***					
RATE LB /HR		0.0000	37544.5781	0.0000	
ACT. RATE BBL/DAY		0.00	3249.67	0.00	
STD. LV RATE BBL/HR		0.00	133.21	0.00	
CP, BTU /LB F		0.0000	0.4767	0.0000	
MOLECULAR WEIGHT....		0.0000	123.7257	0.0000	
ACT. DENS LB /FT3		0.0000	49.3855	0.0000	
STD. API GRAVITY....		0.0000	44.1161	0.0000	
*** DRY BASIS ***					
RATE LB /HR		0.0000	37533.6641	0.0000	
MOLECULAR WEIGHT....		0.0000	123.9373	0.0000	
UOP K		0.0000	11.5026	0.0000	
FLASH POINT, DEG F		0.0000	-77.8037	0.0000	
CRIT. TEMP, F		0.0000	564.1465	0.0000	
CRIT. PRES, PSIA		0.0000	442.3537	0.0000	
*** VAPOR PHASE ***					
RATE LB /HR		4813.1924	0.0000	54.4651	
STD. RATE MM F13/HR		U.UU	U.UUU	0.01	
CP, BTU /LB F		2.1543	0.0000	2.1543	
MOLECULAR WEIGHT....		3.4033	0.0000	3.4033	
VISCOSITY, CP		0.0000	U.UUUU	0.0000	
*** LIQUID PHASE ***					
RATE LB /HR		0.0000	37533.6641	0.0000	
ACT. RATE BBL/DAY		0.00	3249.92	0.00	
CP, BTU /LB F		0.0000	0.4766	0.0000	
MOLECULAR WEIGHT....		0.0000	123.9373	0.0000	
ACT. DENS LB /FT3		0.0000	49.3927	0.0000	
STD. API GRAVITY....		0.0000	44.1260	0.0000	

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STREAM ID..... 21 22
STREAM NAME..... RECYCLE GAS REC COMP DIS
STREAM PHASE..... VAPOR VAPOR
TEMPERATURE, DEG F 120.0000 151.1567
RATE LB PGLS/MR 1400.9648 1400.9609
RATE LB /MR 4785.7979 4785.7539
ENTHALPY MM BTU /MR -2.0511 -1.7423
ENTHALPY BTU /LB -428.5754 -364.0634
MOLECULAR WEIGHT.... 3.4161 3.4161

*** VAPOR PHASE ***
RATE LB /MR 4785.7979 4785.7539
STD.RATE MM FT3/DAY 12.76 12.76
CP, BTU /LB F 2.1464 2.1464
MOLECULAR WEIGHT.... 3.4161 3.4161

*** LIQUID PHASE ***
RATE LB /MR 0.0000 0.0000
ACT.RATE BBL/DAY 0.00 0.00
STD. LV RATE BBL/MR 0.00 0.00
CP, BTU /LB F 0.0000 0.0000
MOLECULAR WEIGHT.... 0.0000 0.0000
ACT.DENS LB /FT3 0.0000 0.0000
STD. API GRAVITY.... 0.0000 0.0000

*** DRY BASIS ***
RATE LB /MR 0.0000 0.0000
MOLECULAR WEIGHT.... 0.0000 0.0000
UOP R 0.0000 0.0000
FLASH POINT, DEG F 0.0000 0.0000
CRIT. TEMP, F 0.0000 0.0000
CRIT. PRES, PSIA 0.0000 0.0000

*** VAPOR PHASE ***
RATE LB /MR 4783.7275 4783.6826
STD.RATE MM FT3/MR 0.53 0.53
CP, BTU /LB F 2.1543 2.1522
MOLECULAR WEIGHT.... 3.4033 3.4033

VISCOSITY, CP 0.0000 0.0000
*** LIQUID PHASE ***
RATE LB /MR 0.0000 0.0000
ACT.RATE BBL/DAY 0.00 0.00
CP, BTU /LB F 0.0000 0.0000
MOLECULAR WEIGHT.... 0.0000 0.0000
ACT.DENS LB /FT3 0.0000 0.0000
STD. API GRAVITY.... 0.0000 0.0000

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	25	26	27	28	29
STREAM NAME	HDC FEED	HCT JP-8	FRAC OVHD	HDT HC VAP	HDT NAPHTHA
STREAM PHASE	LIQUID	LIQUID	VAPOR	VAPOR	LIQUID
TEMPERATURE, DEG F	490.5466	383.7116	305.1799	90.0000	90.0000
PRESSURE, PSIA	35.0000	55.5000	50.0000	40.0000	40.0000
RATE, LB FOLS/HR	57.4040	153.0457	572.4960	65.8279	93.4744
RATE, LB /HR	11660.0053	23950.0117	42544.3281	1111.1794	9150.1523
ENTHALPY MM BTU /HR	1.9736	3.5358	12.3012	0.1937	-0.0266
ENTHALPY BTU /LB	169.2655	147.4333	249.1396	174.3571	-2.8925
MOLECULAR WEIGHT	231.3303	156.4693	74.3132	16.8901	98.6338
*** VAPOR PHASE ***					
RATE, LB /HR	0.0000	0.0000	42544.3281	1111.1794	0.0000
ACT-RATE FT3/SEC	0.00	0.00	25.00	2.68	0.00
STD-RATE MM FT3/DAY	0.00	0.00	5.21	0.60	0.00
CP, BTU /LB F	0.0000	0.0000	0.5071	0.5981	0.0000
MOLECULAR WEIGHT	0.0000	0.0000	74.3138	16.8801	0.0000
ACT-DENS, LB /FT3	0.0000	0.0000	0.4728	0.1151	0.0000
COMPRESSIBILITY (Z)	0.0000	0.0000	0.9576	0.9948	0.0000
*** LIQUID PHASE ***					
RATE, LB /HR	11660.0059	23950.0117	0.0000	0.0000	9190.1523
ACT-RATE BBL/DAY	992.65	2363.61	0.00	0.00	858.17
STD. LV RATE BBL/HR	34.45	92.28	0.00	0.00	35.11
CP, BTU /LB F	0.5736	0.5041	0.0000	0.0000	0.4903
MOLECULAR WEIGHT	231.3308	156.4693	0.0000	0.0000	98.6338
ACT-DENS, LB /FT3	50.2105	43.3135	0.0000	0.0000	45.7763
STD. API GRAVITY	16.7462	38.5351	0.0000	0.0000	57.5816
*** DRY BASIS ***					
RATE, LB /HR	11640.5879	23898.9688	0.0000	0.0000	9188.4551
MOLECULAR WEIGHT	235.9922	159.1016	0.0000	0.0000	98.7155
UOP K	10.5887	11.4572	0.0000	0.0000	11.7568
FLASH POINT, DEG F	228.3018	110.9470	0.0000	0.0000	-18.0632
CRIT. TEMP, PSIA	987.3505	744.5731	0.0000	0.0000	532.0310
CRIT. PRES, PSIA	306.0754	362.9194	0.0000	0.0000	482.2778
*** VAPOR PHASE ***					
RATE, LB /HR	0.0000	0.0000	40607.9063	1090.6182	0.0000
ACT-RATE FT3/SEC	0.00	0.00	20.12	2.64	0.00
STD-RATE MM FT3/DAY	0.00	0.00	0.18	0.02	0.00
CP, BTU /LB F	0.0000	0.0000	0.3117	0.6009	0.0000
MOLECULAR WEIGHT	0.0000	0.0000	87.3287	16.8600	0.0000
ACT-DENS, LB /FT3	0.0000	0.0000	0.3606	0.1149	0.0000
COMPRESSIBILITY (Z)	0.0000	0.0000	0.9490	0.9947	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0100	0.0112	0.0000
*** LIQUID PHASE ***					
RATE, LB /HR	11640.5879	23898.9688	0.0000	0.0000	9188.4551
ACT-RATE BBL/DAY	990.97	2359.59	0.00	0.00	858.05
CP, BTU /LB F	0.5724	0.4031	0.0000	0.0000	0.4902
MOLECULAR WEIGHT	235.9922	159.1016	0.0000	0.0000	98.7155
ACT-DENS, LB /FT3	50.2113	41.2947	0.0000	0.0000	45.7761
STD. API GRAVITY	16.7540	38.5351	0.0000	0.0000	57.5803

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	32	33
STREAM NAME.....	SOUR H2O	REFLUX
STREAM PHASE.....	LIQUID	LIQUID
TEMPERATURE, DEG F	90.0000	90.0000
PRESSURE, PSIA	40.0000	40.0000
RATE LB PCL/HR	105.2445	307.5491
RATE LB /HR	1908.5706	30334.7578
ENTHALPY MM BTU /HR	0.1107	-0.0877
ENTHALPY BTU /LB	58.0136	-2.8925
MOLECULAR WEIGHT....	13.0150	98.6339
*** VAPOR PHASE ***		
RATE LB /HR	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00
STD.RATE MM FT3/DAY	0.00	0.00
CP, BTU /LB F	0.5313	0.0000
MOLECULAR WEIGHT....	18.0150	0.0000
ACT.DENS LB /FT3	0.1273	0.0000
COMPRESSIBILITY (Z).	0.9622	0.0000
*** LIQUID PHASE ***		
RATE LB /HR	1909.5906	30334.7378
ACT.RATE BBL/DAY	131.37	2832.44
STD. LV RATE BBL/HR	5.46	115.89
CP, BTU /LB F	0.9978	0.4904
MOLECULAR WEIGHT....	18.0150	98.6339
ACT.DENS LB /FT3	62.1002	45.7763
STD. API GRAVITY....	10.0635	57.5815
*** DRY BASIS ***		
RATE LB /HR	0.0000	30329.1484
MOLECULAR WEIGHT....	0.0000	98.7156
UOP K	0.0000	11.7568
FLASH POINT, DEG F	0.0000	-18.0632
CRIT. TEMP, F	0.0000	532.0314
CRIT. PRES, PSIA	0.0000	462.2779
*** VAPOR PHASE ***		
RATE LB /HR	0.0000	0.0000
ACT.RATE FT3/SEC	0.00	0.00
STD.RATE MM FT3/HR	0.00	0.00
CP, BTU /LB F	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000
ACT.DENS LB /FT3	0.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000
*** LIQUID PHASE ***		
RATE LB /HR	0.0000	30329.1484
ACT.RATE BBL/DAY	0.00	2932.26
CP, BTU /LB F	0.0000	0.4903
MOLECULAR WEIGHT....	0.0000	98.7156
ACT.DENS LB /FT3	0.0000	45.7741
STD. API GRAVITY....	0.0000	57.5901

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	1	2	3	4	5	6
STREAM NAME	WDT SSC+ LIQUID	MAKEUP H2 VAPOR	HDC RECYCLE LIQUID	NET MAKEUP VAPOR	RX INLET MIXED	RX EFFLUENT MIXED
STREAM PHASE	LIQUID	VAPOR	LIQUID	VAPOR	MIXED	MIXED
TEMPERATURE, DEG F	200.0000	70.0000	510.0000	302.5687	662.0000	683.7269
PRESSURE, PSIA	1240.0000	340.0000	1240.0000	1310.0000	1200.0000	1175.0000
RATE, LB POLS/HR	50.5695	250.0000	43.5297	250.0000	613.2278	1060.3828
RATE, LB /HR	11559.3164	504.3504	11304.9629	504.3504	24256.2656	25780.8398
ENTHALPY MM BTU /HR	0.2596	-0.5197	2.2272	-0.5117	8.5117	12.0444
ENTHALPY MM BTU /LB	22.2675	-1030.5193	157.0121	-240.1482	342.6631	467.1856
MOLECULAR WEIGHT	230.5699	2.0174	259.7073	2.0174	39.5551	24.3128
*** VAPOR PHASE ***						
RATE, LB /HR	0.0000	504.3504	0.0000	504.3504	4392.6396	18838.0352
ACT. RATE, FT3/SEC	0.0000	1.18	0.0000	0.45	1.49	3.02
STD. RATE MM FT3/DAY	0.0000	2.28	0.0000	2.28	4.72	9.34
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4769	1.2568	0.9427
MOLECULAR WEIGHT	0.0000	2.0174	0.0000	2.0174	8.4755	18.3688
ACT. DENS, LB /FT3	0.0000	0.1189	0.0000	0.3098	0.8177	1.7035
COMPRESSIBILITY (Z)	0.0000	1.0145	0.0000	1.0429	1.0334	1.0326
*** LIQUID PHASE ***						
RATE, LB /HR	11659.8164	0.0000	11304.9629	0.0000	19863.6133	6942.8047
ACT. RATE, BBL/DAY	863.65	0.0000	965.10	0.0000	1981.23	761.47
STD. LV RATE BBL/H	34.51	0.0000	34.40	0.0000	60.60	22.06
CP, BTU /LB F	0.4324	0.0000	0.6123	0.0000	0.6722	0.6923
MOLECULAR WEIGHT	230.5699	0.0000	259.7070	0.0000	209.1985	199.2751
ACT. DENS, LB /FT3	57.7093	0.0000	50.0713	0.0000	42.8563	38.9739
STD. API GRAVITY	14.9850	0.0000	19.0979	0.0000	19.5101	25.7666
*** DRY BASIS ***						
RATE, LB /HR	11659.8164	0.0000	11304.9629	0.0000	19863.0898	6942.5146
ACT. RATE, BBL/DAY	863.65	0.0000	965.10	0.0000	1981.23	761.47
STD. LV RATE BBL/H	34.51	0.0000	34.40	0.0000	60.60	22.06
CP, BTU /LB F	0.4324	0.0000	0.6123	0.0000	0.6722	0.6923
MOLECULAR WEIGHT	230.5699	0.0000	259.7070	0.0000	209.1985	199.2751
ACT. DENS, LB /FT3	57.7093	0.0000	50.0713	0.0000	42.8563	38.9739
STD. API GRAVITY	14.9850	0.0000	19.0979	0.0000	19.5101	25.7666
*** VAPOR PHASE ***						
RATE, LB /HR	0.0000	504.3504	0.0000	504.3504	4385.6611	18816.2578
ACT. RATE, FT3/SEC	0.0000	1.18	0.0000	0.45	1.49	3.02
STD. RATE MM FT3/DAY	0.0000	2.28	0.0000	2.28	4.72	9.34
CP, BTU /LB F	0.0000	3.4201	0.0000	3.4769	1.2568	0.9427
MOLECULAR WEIGHT	0.0000	2.0174	0.0000	2.0174	8.4755	18.3688
ACT. DENS, LB /FT3	0.0000	0.1189	0.0000	0.3098	0.8177	1.7035
COMPRESSIBILITY (Z)	0.0000	1.0145	0.0000	1.0429	1.0334	1.0326
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE, LB /HR	11659.8164	0.0000	11304.9629	0.0000	19863.0898	6942.5146
ACT. RATE, BBL/DAY	863.65	0.0000	965.10	0.0000	1981.23	761.47
STD. LV RATE BBL/H	34.51	0.0000	34.40	0.0000	60.60	22.06
CP, BTU /LB F	0.4324	0.0000	0.6123	0.0000	0.6722	0.6923
MOLECULAR WEIGHT	230.5699	0.0000	259.7070	0.0000	209.1985	199.2751
ACT. DENS, LB /FT3	57.7093	0.0000	50.0713	0.0000	42.8563	38.9739
STD. API GRAVITY	14.9850	0.0000	19.0979	0.0000	19.5101	25.7666

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STREAM ID.	7	8	9	10	11	12
STREAM NAME	HOT SEP VAP	EFF FPCW E-2	OUTLET COLD SEP	RECYLE GAS	PURGE GAS	
STREAM PHASE	VAPOR	MIXED	MIXED	VAPOR	VAPOR	
TEMPERATURE, DEG F	663.6332	584.8690	379.3525	365.2185	120.0000	120.0000
PRESSURE, PSIA	1170.0000	1145.0000	1120.0000	1095.0000	1095.0000	1095.0000
RATE, LB PCL/HR	1025.7037	1025.7036	1025.7037	1039.5911	791.5305	115.9933
RATE, LB /HR	1362.0391	18562.0391	18662.0352	19112.0352	2315.1357	339.2587
ENTHALPY MK BTU /HR	7.7359	7.6068	3.5190	3.5429	-1.2087	-0.1771
ENTHALPY BTU /LB	515.1650	403.2874	186.5671	185.3755	-522.0969	-522.0966
MOLECULAR WEIGHT.....	19.3894	19.3894	18.3494	18.3944	2.9248	2.9248
*** VAPOR PHASE ***						
RATE, LB /HR	18862.0391	13840.3379	7340.9643	7205.7803	2315.1357	339.2587
ACT-RATE, FT3/SEC	3.03	2.81	2.20	2.21	1.30	0.19
STD. RATE MM FT3/DAY	9.34	9.10	8.69	8.72	7.21	1.06
CP, BTU /LB F	0.9423	1.0164	1.2813	1.2866	2.4858	2.4858
MOLECULAR WEIGHT.....	18.3894	13.8471	7.6972	7.4761	2.9248	2.9248
ACT-DENS, LB /FT3	1.69E5	1.3599	0.9251	0.9067	0.4932	0.4932
COMPRESSIBILITY (Z).	1.0325	1.0326	1.0350	1.0340	1.0439	1.0439
*** LIQUID PHASE ***						
RATE, LB /HR	0.0000	5021.7061	11521.1738	11906.2520	0.0000	0.0000
ACT-RATE, GBL/DAY	0.00	522.39	1139.93	1172.41	0.00	0.00
STD. LV RATE GBL/HR	0.00	16.30	40.01	41.50	0.00	0.00
CP, BTU /LB F	0.0000	0.6770	0.6182	0.6133	0.0000	0.0000
MOLECULAR WEIGHT.....	0.0000	191.7433	160.0094	157.2109	0.0000	0.0000
ACT-DENS, LB /FT3	0.0000	41.0911	43.2028	43.4098	0.0000	0.0000
STD. API GRAVITY.....	0.0000	29.1072	40.3978	41.0296	0.0000	0.0000
*** DRY BASIS ***						
RATE, LB /HR	0.0000	5021.4668	11520.3184	11894.9570	0.0000	0.0000
ACT-RATE, GBL/DAY	0.0000	191.8315	160.1033	158.3731	0.0000	0.0000
UOP K	0.0000	11.2991	11.8781	11.6975	0.0000	0.0000
FLASH POINT, DEG F	0.0000	0.3040	-38.4711	-39.4843	0.0000	0.0000
CRIT. TEMP, F	0.0000	710.1705	634.0975	631.0258	0.0000	0.0000
CRIT. PRES, PSIA	0.0000	281.3445	318.6971	321.7490	0.0000	0.0000
*** VAPOR PHASE ***						
RATE, LB /HR	18840.2572	13518.7949	7319.9404	6945.2979	2293.0664	336.0247
ACT-RATE, FT3/SEC	3.03	2.80	2.20	2.18	1.30	0.19
STD. RATE MM FT3/DAY	0.39	0.38	0.36	0.36	0.30	0.04
CP, BTU /LB F	0.9429	1.0172	1.2835	1.3173	2.5034	2.5034
MOLECULAR WEIGHT.....	18.3894	13.8421	7.6847	7.3156	2.9014	2.9014
ACT-DENS, LB /FT3	1.69E5	1.3593	0.9235	0.8867	0.4892	0.4892
COMPRESSIBILITY (Z).	1.0325	1.0326	1.0351	1.0346	1.0440	1.0440
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***						
RATE, LB /HR	0.0000	5021.4569	11520.3184	11894.9570	0.0000	0.0000
ACT-RATE, GBL/DAY	0.00	522.37	1139.86	1171.53	0.00	0.00
CP, BTU /LB F	0.0000	0.6769	0.6182	0.6129	0.0000	0.0000
MOLECULAR WEIGHT.....	0.0000	191.5315	160.1033	155.3731	0.0000	0.0000
ACT-DENS, LB /FT3	0.0000	41.0910	43.2022	41.4011	0.0000	0.0000
STD. API GRAVITY.....	0.0000	29.1072	40.3978	41.0296	0.0000	0.0000

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	13	14	15	16	17
STREAM NAME	HOT SEP LIG	COLD SEP LIG	REC COMP OIS	H2O WASH	SOUR H2O
STREAM PHASE	LIQUID	LIQUID	VAPOR	LIQUID	LIQUID
TEMPERATURE, DEG F	983.6332	120.0000	153.7265	125.0000	120.0000
PRESSURE, PSIA	1170.0000	1095.0000	1245.0000	1095.0000	1095.0000
RATE, LB FOLS/HR	34.6753	113.5703	731.5386	13.8773	13.4675
RATE, LB /HR	6918.8027	16215.1074	2315.1602	249.9999	242.5315
ENTHALPY MM BTU /HR	2.3081	0.2972	-1.0187	0.0239	0.0063
ENTHALPY BTU /LB	333.5910	13.3297	-439.9921	95.7615	25.9441
MOLECULAR WEIGHT	199.5084	135.7552	2.6248	18.0150	18.0387
*** VAPOR PHASE ***					
RATE, LB /HR	0.0000	0.0000	2315.1602	0.0000	0.0000
ACT. RATE, FT3/SEC	0.00	0.00	1.20	0.00	0.00
STD. RATE MM FT3/DAY	0.00	0.00	7.21	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	2.4957	0.0000	0.0000
MOLECULAR WEIGHT	0.0000	0.0000	2.9248	0.0000	0.0000
ACT. DENS, LB /FT3	0.0000	0.0000	0.5358	0.0000	0.0000
COMPRESSIBILITY (Z)	0.0000	0.0000	1.3451	0.0000	0.0000
*** LIQUID PHASE ***					
RATE, LB /HR	6918.8027	16215.1074	0.0000	249.9999	242.5315
ACT. RATE, BBL/DAY	752.47	1446.55	0.00	17.34	16.82
STD. LV RATE BBL/HH	21.98	59.02	0.00	0.72	0.69
CP, BTU /LB F	0.6924	0.4958	0.0000	0.9929	1.1699
MOLECULAR WEIGHT	199.5084	135.7552	0.0000	18.0150	18.0087
ACT. DENS, LB /FT3	39.5990	47.9157	0.0000	61.6306	61.6314
STD. API GRAVITY	25.7100	42.6534	0.0000	10.0635	10.2821
*** DRY BASIS ***					
RATE, LB /HR	6918.5146	16210.8398	0.0000	0.0000	0.3195
MOLECULAR WEIGHT	199.5922	136.9928	0.0000	0.0000	14.2259
UOP K	11.1456	11.5915	0.0000	0.0000	13.0285
FLASH POINT, DEG F	19.0925	-66.1856	0.0000	0.0000	1.9601
CRIT. TEMP, F	722.4723	584.6508	0.0000	0.0000	6.5419
CRIT. PRES, PSIA	271.2542	367.0006	0.0000	0.0000	1028.8279
*** VAPOR PHASE ***					
RATE, LB /HR	0.0000	0.0000	2253.0903	0.0000	0.0000
ACT. RATE, FT3/SEC	0.00	0.00	1.20	0.00	0.00
STD. RATE MM FT3/HH	0.00	0.00	7.21	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	2.5153	0.0000	0.0000
MOLECULAR WEIGHT	0.0000	0.0000	2.9014	0.0000	0.0000
ACT. DENS, LB /FT3	0.0000	0.0000	0.5315	0.0000	0.0000
COMPRESSIBILITY (Z)	0.0000	0.0000	1.0492	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0000	0.0000	0.0000
*** LIQUID PHASE ***					
RATE, LB /HR	6918.5146	16210.8398	0.0000	0.0000	0.3195
ACT. RATE, BBL/DAY	758.43	1446.55	0.00	0.00	0.04
CP, BTU /LB F	0.6923	0.4956	0.0000	0.0000	0.7504
MOLECULAR WEIGHT	199.5922	135.9928	0.0000	0.0000	14.2259
ACT. DENS, LB /FT3	39.5990	47.9129	0.0000	0.0000	31.7042
STD. API GRAVITY	25.7100	42.6534	0.0000	0.0000	175.9743

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	25	26	27	28	30	31
STREAM NAME	HDC NAPHTHA	HDC JP-8	FRAC CUMD	HDC HC VAP	FRAC STR STM JP-8	STR STM VAPOR
STREAM PHASE	LIQUID	LIQUID	VAPOR	VAPOR	VAPOR	VAPOR
TEMPERATURE, DEG F	110.0000	354.9146	291.1038	110.0000	318.6042	311.6934
PRESSURE, PSIA	150.0000	55.0000	60.0000	50.0000	67.5000	55.0000
RATE LB MOLS/MR	53.2450	34.3234	222.2137	16.2345	48.2931	13.2651
RATE LB /LB F	5649.8533	5929.9930	15413.7305	295.7139	865.9995	274.9995
ENTHALPY MM BTU /MR	0.0985	0.8636	4.8493	0.0352	1.0352	0.3272
ENTHALPY BTU /LB	17.4365	145.6345	314.6068	121.5405	1189.8687	1189.8687
MOLECULAR WEIGHT....	96.9982	145.2557	69.5644	18.1927	18.0150	18.0150
*** VAPOR PHASE ***						
RATE LB /MR	0.0000	C.0000	15413.7305	295.7139	865.9995	274.9995
ACT. RATE FT3/SEC	0.0000	0.00	7.91	0.35	1.60	0.62
STD. RATE MM FT3/DAY	0.00	0.00	2.02	0.15	0.44	0.14
CP, BTU /LB F	0.0000	0.0000	0.5339	0.6564	0.5462	0.5305
MOLECULAR WEIGHT....	0.0000	0.0000	69.5644	18.1927	18.0150	18.0150
ACT. DENS LB /FT3	0.0000	0.0000	0.5412	0.1434	0.1513	0.1236
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9545	0.9940	0.9623	0.9685
*** LIQUID PHASE ***						
RATE LB /MR	5649.8533	5929.9980	0.0000	0.0000	C.0000	0.0000
ACT. RATE BBL/DAY	571.89	606.46	0.00	0.00	0.00	0.00
STD. LV RATE BBL/MR	23.07	21.40	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.5374	0.6237	0.0000	0.0000	C.0000	0.0000
MOLECULAR WEIGHT....	96.9982	163.2557	0.0000	0.0000	C.0000	0.0000
ACT. DENS LB /FT3	42.2281	41.7968	0.0000	0.0000	0.0000	0.0000
STD. API GRAVITY....	70.5799	47.0949	0.0000	0.0000	C.0000	0.0000
*** DRY BASIS ***						
RATE LB /MR	5647.9697	5916.9121	0.0000	0.0000	C.0000	0.0000
MOLECULAR WEIGHT....	97.1252	166.2198	0.0000	0.0000	C.0000	0.0000
UOP K	12.4155	12.0317	0.0000	0.0000	C.0000	0.0000
FLASH POINT, DEG F	-40.8827	96.6094	0.0000	0.0000	C.0000	0.0000
CRIT. TEMP, F	490.8982	718.5243	0.0000	0.0000	C.0000	0.0000
CRIT. PRES, PSIA	437.1967	322.2452	0.0000	0.0000	C.0000	0.0000
*** VAPOR PHASE ***						
RATE LB /MR	0.0000	0.0000	14296.7852	288.2875	0.0000	0.0000
ACT. RATE FT3/SEC	0.00	0.00	5.62	0.34	0.00	0.00
STD. RATE MM FT3/MR	0.00	0.00	0.06	0.01	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.5428	0.6616	0.0000	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	89.2269	19.1973	C.0000	0.0000
ACT. DENS LB /FT3	0.0000	0.0000	0.7068	0.1494	C.0000	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9403	0.9960	C.0000	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0097	0.0122	C.0000	0.0000
*** LIQUID PHASE ***						
RATE LB /MR	5647.9697	5916.9121	0.0000	0.0000	C.0000	0.0000
ACT. RATE BBL/DAY	571.77	603.45	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.5373	0.5231	0.0000	0.0000	C.0000	0.0000
MOLECULAR WEIGHT....	97.1252	166.2198	0.0000	0.0000	C.0000	0.0000
ACT. DENS LB /FT3	42.2241	41.7741	0.0000	0.0000	C.0000	0.0000
STD. API GRAVITY....	72.5553	47.1753	0.0000	0.0000	C.0000	0.0000

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID. 12
STREAM NAME SOUR H2O
STREAM PHASE LIQUID
TEMPERATURE, DEG F 110.0000
PRESSURE, PSIA 50.0000
RATE, LB MOLS/HR 51.4552
RATE, LB /HR 1107.3677
ENTHALPY MM BTU /HR 0.0663
ENTHALPY BTU /LB 77.9702
MOLECULAR WEIGHT.... 12.0150

*** VAPOR PHASE ***
RATE, LB /HR 0.0000
ACT-RATE, FT3/SEC 0.00
STD-RATE MM FT3/DAY 0.00
CP, BTU /LB F 0.0000
MOLECULAR WEIGHT.... 0.0000
ACT-DENS, LB /FT3 0.0000
COMPRESSIBILITY (Z). 0.0000

*** LIQUID PHASE ***
RATE, LB /HR 1107.3677
ACT-RATE, SBL/DAY 76.53
STD. LV RATE SBL/HR 3.17
CP, BTU /LB F 0.9978
MOLECULAR WEIGHT.... 18.0150
ACT-DENS, LB /FT3 61.4534
STD. API GRAVITY.... 10.0635

*** DRY BASIS ***
RATE, LB /HR 0.0000
MOLECULAR WEIGHT.... 0.0000
UOP K 0.0000
FLASH POINT, DEG F 0.0000
CRIT. TEMP, F 0.0000
CRIT. PRES, PSIA 0.0000

*** VAPOR PHASE ***
RATE, LB /HR 0.0000
ACT-RATE, FT3/SEC 0.00
STD-RATE MM FT3/HR 0.00
CP, BTU /LB F 0.0000
MOLECULAR WEIGHT.... 0.0000
ACT-DENS, LB /FT3 0.0000
COMPRESSIBILITY (Z). 0.0000
VISCOSITY, CP 0.0000

*** LIQUID PHASE ***
RATE, LB /HR 0.0000
ACT-RATE, SBL/DAY 0.00
CP, BTU /LB F 0.0000
MOLECULAR WEIGHT.... 0.0000
ACT-DENS, LB /FT3 0.0000
STD. API GRAVITY.... 0.0000

33
REFLUX
LIQUID

110.0000
50.0000
56.3171
9369.8837
0.1463
17.4235
96.4567

9369.8887
848.50
34.17
0.5373
96.9667
42.1456
70.5915

8367.3945
97.0937
12.4155
-40.9109
490.7809
437.2733

0.0000
0.00
0.00
0.0000
0.0000
0.0000
0.0000

8367.3945
54.553
5.5372
47.5977
42.1515
71.5344

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PROBLEM C3U200

TM
PROCESS
SOLUTION

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REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	40	41	42	43	44
STREAM NAME	NAPHTHA FEED	STAG OFFGAS	REFLUX	NAPHTHA PROD	REBOILER
STREAM PHASE	LIQUID	VAPOR	LIQUID	LIQUID	LIQUID
TEMPERATURE, DEG F	97.8288	109.2971	109.2971	372.6451	314.1483
PRESSURE, LB /SQ IN	144.0000	125.0000	125.0000	142.0000	141.5000
RATE, LB /HR	150.8883	7.5061	18.7648	143.2816	208.7294
RATE, BTU /HR	14776.7215	319.9324	1045.5327	14455.0059	29079.9414
ENTHALPY, MM BTU /HR	0.0721	0.0679	0.0451	2.4428	3.9098
ENTHALPY, BTU /LB	4.8822	212.3046	43.1618	168.9918	134.4507
MOLECULAR WEIGHT	97.9319	42.6228	55.7178	100.8853	94.1923
*** VAPOR PHASE ***					
RATE, LB /HR	0.0000	319.9324	0.0000	0.0000	0.0000
ACT. RATE, FT3 /SEC	0.00	0.09	0.00	0.00	0.00
STD. RATE, MM FT3 /DAY	0.00	0.07	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.4313	0.0000	0.0000	0.0000
MOLECULAR WEIGHT	0.0000	42.6228	0.0000	0.0000	0.0000
ACT. DENS, LB /FT3	0.0000	0.9854	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z)	0.0000	0.8856	0.0000	0.0000	0.0000
*** LIQUID PHASE ***					
RATE, LB /HR	14776.7715	0.0000	1045.5327	14455.0059	29079.9414
ACT. RATE, BBL /DAY	1422.75	0.00	133.44	1793.74	3478.13
STD. LV RATE, BBL /HR	57.54	0.00	5.22	56.26	115.13
CP, BTU /LB F	0.5085	0.0000	0.7083	0.7027	0.6870
MOLECULAR WEIGHT	97.9319	0.0000	55.7178	100.8853	94.1923
ACT. DENS, LB /FT3	44.3959	0.0000	33.4915	34.4469	35.7388
STD. API GRAVITY	62.5768	0.0000	115.3759	61.1234	64.4487
*** DRY BASIS ***					
RATE, LB /HR	14773.3887	0.0000	1044.9988	14454.7793	29075.9336
MOLECULAR WEIGHT	99.0315	0.0000	55.7775	100.8926	94.2474
UOP K	12.0088	0.0000	13.6353	11.9656	12.0326
FLASH POINT, DEG F	-22.1453	0.0000	-115.0079	-7.3064	-38.9144
CRIT. TEMP, F	515.6722	0.0000	284.2679	530.8590	500.6957
CRIT. PRES, PSIA	465.1137	0.0000	587.5289	448.3092	468.9988
*** VAPOR PHASE ***					
RATE, LB /HR	0.0000	312.5894	0.0000	0.0000	0.0000
ACT. RATE, FT3 /SEC	0.00	0.09	0.00	0.00	0.00
STD. RATE, MM FT3 /DAY	0.00	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.4818	0.0000	0.0000	0.0000
MOLECULAR WEIGHT	0.0000	42.8699	0.0000	0.0000	0.0000
ACT. DENS, LB /FT3	0.0000	0.9923	0.0000	0.0000	0.0000
COMPRESSIBILITY (Z)	0.0000	0.8845	0.0000	0.0000	0.0000
VISCOSITY, CP	0.0000	0.0091	0.0000	0.0000	0.0000
*** LIQUID PHASE ***					
RATE, LB /HR	14773.3557	0.0000	1044.9988	14454.7793	29075.9336
ACT. RATE, BBL /DAY	1422.52	0.00	133.41	1793.72	3477.83
CP, BTU /LB F	0.5054	0.0000	0.7081	0.7027	0.6870
MOLECULAR WEIGHT	98.0315	0.0000	55.7775	100.8926	94.2474
ACT. DENS, LB /FT3	44.3423	0.0000	33.4436	34.4467	35.7369
STD. API GRAVITY	62.5768	0.0000	115.3759	61.1242	64.4562

APPENDIX E

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 4
PROFITABLE JP-8 PRODUCTION
SUBTASKS 1.2 & 1.3
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
DATE - JAN. 30, 1988

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1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to produce JP-8 type aviation turbine fuel and chemical byproducts to maximize profit from Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- . Tar Oil byproduct stream (47620 #/hr, 3182 BPSD) is charged to the hydrotreater (Area 100).
- . The hydrotreater is a 3 stage expanded bed type process which removes 99% + of the sulfur, nitrogen, and oxygen compounds and begins the conversion of 550°F+ material. The hydrotreater adds a large quantity of hydrogen to the feed (4100 SCF/bbl) which results in a high heat of reaction. An expanded bed type reactor was chosen to both control and utilize the heat of reaction. Three stages were used to both control the temperature rise as well as to obtain the high efficiency associated with staging a back-mixed reactor.
- . The hydrotreater produces 6 streams:
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H₂ and CH₄.
 - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.
 - Unstabilized naphtha which is sent to the combined naphtha stabilizer in the hydrocracker (area 200). After stabilization, to control vapor pressure, the naphtha is sent to storage and gasoline blending.
 - JP-8 turbine fuel which is combined with JP-4 produced in the hydrocracker (area 200) and sent to storage.
 - 550°F+ unconverted bottoms product which is sent to the hydrocracker (area 200).
 - Wastewater containing, NH₄OH and NH₄HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H₂S and NH₃.

- Approximately 950 #/day of spent catalyst which is shipped to a catalyst reclaimer in the same drums that the catalyst is received in.
- . The 550⁰F+ unconverted stream from the expanded bed hydrotreater (area 100) is charged to the fixed bed hydrocracker (area 200). The hydrocracker converts this material to naphtha and JP-8 turbine fuel. For this service a 5 stage unit was chosen with 65% conversion per pass. This unit also includes a naphtha stabilizer which stabilizes both the naphtha produced in the hydrotreater and hydrocracker.
- . The hydrocracker produces 4 streams in addition to JP-8
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit of the SNG plant for recovery of the H₂ and CH₄.
 - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.
 - Stabilized naphtha which is sent to storage and gasoline blending.
 - A small sour water stream which is sent to the PHOSAM unit in the SNG plant or alternatively used as part of the injection water to the hydrotreating plant.
- . Hydrogen make-up for the Hydrotreater, the Hydrocracker and the Naphtha Hydrotreater is supplied from a PSA Hydrogen Unit (Area 300). High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining purge gas is available a low pressure (5 psig) which has a fuel value of about 565 BTU/ft³. This H₂, CO & CH₄ rich gas is recompressed into the methanation Unit of the SNG plant.
- . The crude naphtha byproduct stream (8738#/hr, 725 BPSD) is charged to the distillation and hydrotreating unit (Area 600).
- . The distillation removes the material boiling below 160⁰F, which is sent to the SNG Plant fuel pool, and produces a bottoms product which is charged to the hydrotreater.
- . The fixed bed hydrotreater is a single bed reactor which removes 99% + of the sulfur, nitrogen, and oxygen compounds. Hydrogen is added to the feed at the rate of 430 SCF/bbl.

1.1 Overall Process Description - cont'd

- . The naphtha hydrotreater produces 4 streams:
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H_2 and CH_4 .
 - Naphtha which is stabilized to control vapor pressure, and the sent to the aromatics recovery unit (Area 700).
 - A low pressure off gas which is sent to the Stretford unit in the SNG plant.
 - Wastewater containing, NH_4OH and NH_4HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H_2S and NH_3 .
- . The hydrotreated naphtha is charged to the extraction section of the Aromatics Recovery Unit (Area 700) where it is contacted with a solvent to extract the aromatic components from the stream. The raffinate is sent to storage and gasoline blending while the solvent is recovered from the aromatic extract. The aromatic extract is then sent to fractionation to produce the BTX products.
- . Five streams are produced in the ARU plant.
 - A hydrocarbon gasoline blending stock which is sent to storage and gasoline blending.
 - A small process water stream which is sent to the waste treatment plant in the SNG Plant.
 - Three product streams Benzene, Toluene & Xylene which are sent to storage.
- . The crude phenol byproduct stream (14490 #/hr, 936 BPSD), is feed to the dual solvent phenol extraction unit (Area 800).

1.1 Overall Process Description - cont'd

- . Distillation removes approx. 85% of the phenol which is further distilled to remove light ends and then reflashed over sulfuric acid producing a 99.8% pure product.
- . The remainder of the stream (a cresylic acid mixture) is flash distilled over a 3 wt.% concentrated sulfuric acid mixture to remove pyridine type substances.
- . The acid tar produced is water washed and mixed with light oil and sent to fuel.
- . The remaining cresol/xlenol mixture is double solvent extracted to remove neutral hydrocarbons. The resulting crude cresylic acid is dried and sent either to storage or distillation (Area 900).
- . Streams produced in the phenol extraction unit are:
 - Phenol product sent to storage
 - Crude Cresylic Acid sent to distillation (Area 900) or storage.
 - Wash Water sent to Water Treatment in the SNG Plant.
 - Waste Water sent to the Phenosolvan unit in the SNG Plant.
 - Neutral Oil sent to storage and fuel for the SNG Plant boilers.
- . The Crude Cresylic Acid is progressively distilled (Area 900) to separate the cresols and xylenols. No attempt has been made to remove the guaiacol from the product streams.
- . Streams produced in the crude cresylic acid distillation unit are:
 - o-Cresol product which is sent to storage.
 - m,p-Cresol product which is sent to storage.
 - Xlenol product which is sent to storage.
 - A heavy distillate which is combined with neutral oil in Area 800.

1.1 Overall Process Description - cont'd

- A crude phenol stream which is recycled to the Area 800.
- A small water stream which is sent to Area 800 for tar acid washing.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas, neutral oil and 160°F-distillate produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

936 BPSD of Crude Phenol
725 BPSD of Crude Naphtha
3182 BPSD of Tar Oil
4361 BPSD of #6 Fuel Oil
10.76 MMSCFD equivalent SNG product loss due to the syn gas feed to the PSA unit.

Products

2552 BPSD of JP-8 turbine fuel
1276 BPSD of 300°F - Naphtha for gasoline blending
317 BPSD of Phenol
56 BPSD of o-Cresol
131 BPSD of m,p-Cresol
75 BPSD of Xylenols
312 BPSD of Neutral Oil for Fuel
202 BPSD of 160°F - Distillate for Fuel
46 BPSD of Gasoline Blending Stock
315 BPSD of Benzene
112 BPSD of Toluene
15 BPSD of Xylene
7.02 MMSCFD equivalent SNG product credit due to HDT, HDC & PSA purge gas reinjection into SNG plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	4361 BPSD
SNG Equivalent	
of Syn Gas & Purge Gas	3.73 MM SCFD
Power	6160 kW
Cooling Water	6050 GPM (30°F rise)
Process Water	24 GPM

In addition the process utilizes steam as summarized below which was debited against boiler requirements.

HP Steam Import	58,200 #/H
MP Steam Import	8,900 #/H
LP Steam Export	6,900 #/H
Condensate Return	67,100 #/H

Figure 1 Case 4: Profitable JP-8

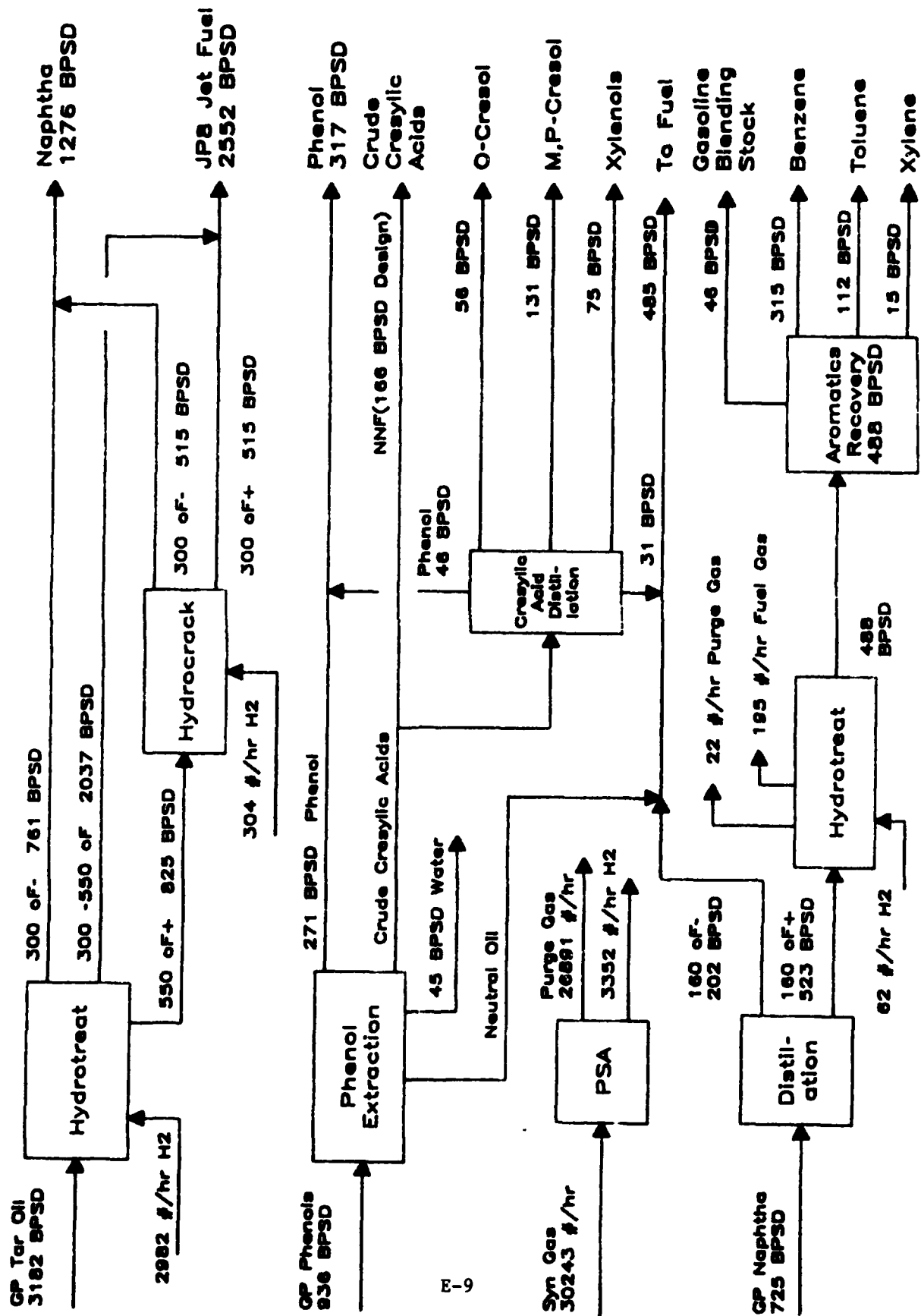


Table 1.1 Great Plains Case 4: Profitable JPB Production

=====				
Tar Oil Feed====>	47620	#/hr	3182	BPSD
Phenol Feed====>	14490	#/hr	936	BPSD
Cr'd Naphtha Feed=>	8738	#/hr	725	BPSD
Naphtha Product==>	13588	#/hr	1276	BPSD
JPB Product====>	30631	#/hr	2552	BPSD
Phenol Product==>	4925	#/hr	317	BPSD
o-Cresol Prod====>	845	#/hr	56	BPSD
m,p-Cresol Prod==>	1974	#/hr	131	BPSD
Xylenols Prod====>	1070	#/hr	75	BPSD
Gasoline Stock==>	481	#/hr	46	BPSD
Benzene Prod====>	4060	#/hr	315	BPSD
Toluene Prod====>	1425	#/hr	112	BPSD
Xylene Prod====>	188	#/hr	15	BPSD
SNP Product Loss=>	6411	#/hr	3.7	MMSCFD
Fuel Oil Makeup==>	60370	#/hr	4361	BPSD

Expanded Bed Hydrotreater

=====					
Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD

Feeds					
H2	6.26		2982	1479.2	
Tar Oil	100.00	1.0268	47620		3182

Total	106.26		50602		
Products					
Purge Gas	0.10		48	14.9	
Fuel Gas	1.87		891	43.9	
Naphtha	17.70	0.7600	8430		761
JP-B	51.87	0.8320	24701		2037
550 oF+	24.49	0.9700	11663		825
H2O in SW	8.73		4159	231.1	
H2S in SW	0.43		205	6.0	
NH3 in SW	1.06		505	29.7	

Total	106.26		50602		3623

Fixed Bed Hydrocracker

=====					
Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD

Feeds					
H2	2.61		304	150.8	
500oF+	100.00	0.9700	11663		825

Total	102.61		11967		
Products					
Purge Gas	1.48		173	58.6	
Fuel Gas	4.27		498	27.5	
Naphtha	46.00	0.7148	5365		515
JP-B	50.84	0.7900	5930		515
H2S in SW	0.003		0.4	0.01	
NH3 in SW	0.003		0.4	0.02	

Total	102.60		11967		1030

Naphtha Stabilizer

Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD
HDT Nap	61.11	0.7600	8430		761
HCR Nap	38.89	0.7148	5365		515
Stab Nap	98.50	0.7306	13588		1276
Fuel Gas	1.50		207	4.8	

PSA Hydrogen Recovery Unit(86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	8.63	5.55	903.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	1931.1	569.0	45.2	495.6	9.6	5.8	3056.2
Prod. H2	1660.7	0.2	0.0	0.0	0.0	0.0	1660.9
Purge Gas	270.4	568.8	45.2	495.6	9.6	5.8	1395.3
#/hr							
Feed Gas	3893	15935	1990	7950	289	186	30243
Prod. H2	3348	4	0	0	0	0	3352
Purge Gas	545	15931	1990	7950	289	186	26891

Crude Naphtha
Distillation

	Wt %	Gravity	#/hr	BPSD
Feed Naphtha	100.00	0.8269	8738	725
Prod 160 oF-	24.77	0.7350	2164	202
Prod 160 oF+	75.23	0.8627	6574	523

Naphtha Hydrotrater

=====

Component	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feed 160 oF+	100.00	0.8627	6574		523
Feed Hydrogen	0.94		62	30.8	
Feed Total	100.94		6636		523
Products					
Purge Gas	0.33		22	6.8	
Fuel Gas	2.97		195	10.8	
HDT Naphtha	93.61	0.8650	6154		488
H ₂ O in SW	1.96		129		
H ₂ S in SW	1.76		116		
NH ₃ in SW	0.30		20		
Total Products	100.94		6636		488

Aromatics Recovery

=====

Component	Wt %	Grav	#/hr	BPSD
Feed HDT Naphtha	100.00	0.8650	6154	488
Products				
Raffinate	7.82	0.7175	481	46
Benzene	65.97	0.8844	4060	315
Toluene	23.16	0.8718	1425	112
Xylene	3.05	0.8729	188	15
Total Products	100		6154	488

Phenol Extraction

=====

Component	Wt %	#/hr	Grav	BPSD
Feeds				
Crude Phenol	100.00	14490	1.0621	936
Sulfuric Acid	1.97	285	1.8300	11
Total Feed	101.97	14775		947
Products				
Cr. Cresylic Acid	35.13	5090	1.0290	339
Phenol	29.09	4215	1.0661	271
Neutral Oil	30.68	4445	1.0860	281
Acidic Waste Water	7.07	1025	1.2558	56
Total Products	101.97	14775		947

Cresylic Acid Distillation

Component	Wt %	#/hr	Grav	BPSD
Cr. Cresylic Acid	100.00	5090	1.0290	339
Products				
Phenol	13.95	710	1.0661	46
o-Cresol	16.60	845	1.0350	56
m,p-Cresol	38.78	1974	1.0340	131
Xylenols	21.02	1070	0.9750	75
Heavies	9.65	491	1.0800	31
Total	100.00	5090		339

Fuel Gas Generated in Hydrotreating and Hydrocracking

Component	#/hr	#Mol/hr	MMBTU/hr
HDTR FG Produced	891	43.9	16.0
HCR FG Produced	498	27.5	9.0
Stabilizer FG	207	4.8	3.7
Naphtha Hdtr FG	195	10.8	3.5
Total Fuel Gas	1596	76.3	28.7

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	545	270.4	324	33.2
CO	15931	568.8	321	69.2
CO2	1990	45.2	0	0.0
C1	7950	495.6	1010	189.7
C2	289	9.6	1769	6.5
N2+Ar	186	5.8	0	0.0
Total	26891	1395.3	565	298.6

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft3	BPSD
Tar Oil	-47620	17000	-809.5			-3182
Crude Phenol	-14490	13070	-189.4			-936
Crude Naphtha	-8738	18500	-161.7			-725
Fuel Gas	1596	18000	28.7	0.7	994	
160 oF- distillate	2164	17400	37.7			202
Neutral Oil	4936	15000	74.0			312
Import Steam	-60200	1000	-60.2			
Fuel Oil to Boiler	60020	18000	1080.4			4336
Total	-62332		0.0	0.7		7
Fuel Oil to Process Heaters	350	18000	6.3			25

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	10.76	73350
SNG Credit for PSA Purge gas	6.74	33488
SNG Credit for Hdtrs purge gas	0.29	1929
Total SNG Production Loss	3.73	

2.0 PROCESS DESCRIPTION

2.1 Hydrotreater (Area 100)

For a description of the Hydrotreater process see Case 3 Section 2.1.

2.2 Hydrocracker (Area 200)

For a description of the Hydrocracker process see Case 3 Section 2.2.

2.3 PSA Hydrogen Unit & Recompression (Area 300)

2.3.1 Hydrogen for both the hydrotreater and the hydrocracker will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig
Temp.	65 °F
Composition	mol%

H2	63.19
CO	18.61
CO2	1.48
CH4	16.21
C2H6	0.31
COS, H2S, CS2	< 0.01
N2 + Ar	0.19
H2O	< 0.01

The PSA unit selectively absorbs all components except H2 and produces a 99.99% vol. purity stream at about 345 psig and 80°F. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100°F
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

At the conditions given a 10 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

2.3 PSA Hydrogen Unit & Recompression (Area 300) - cont'd

2.3.1 Cont'd

The system uses 10 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continuously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-40301 presents a schematic of a Union Carbide Polybed PSA unit.

2.3.2 The purge gas is recompressed to 375 psia and sent to the methanation unit of the SNG plant.

2.4 Phenol Stream (Areas 800 and 900)

For a description of the Phenol Extraction (Area 800) and Cresylic Acid Distillation (Area 900) Units see Case 7 Section 2.1.

2.5 Naphtha Stream (Area 600 and 700)

For a description of the Naphtha Distillation and Hydrotreating Unit (Area 600) and the Aromatics Recovery Unit (Area 700) see Case 7 Section 2.2.

PAGE 2-3

TYPICAL ARRANGEMENT

NUMBER OF ABSORBERS FOR THIS CASE = 10

[illegible]

AMOCO/DOE
 GREAT PLAINS GASIFICATION PLANT
 JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.1 Equipment List

CASE 4 - PROFITABLE JP-8

AREA 100 - HYDROTREATER

TAG. NO. DESCRIPTION

See Case 3 Area 100

AREA 200 - HYDROCRACKER

See Case 3 Area 200

AREA 300 - PSA HYDROGEN UNIT & RECOMPRESSION

FA-301 Purge Gas Surge Drum

GB-301 Purge Gas Compressor

PA-301 PSA Hydrogen Unit Package

AREA 400 - STORAGE AREA

FB-401 Jet Fuel Storage Tank

FB-402 Naphtha Storage Tank

FB-403 Fuel Oil Storage Tank

FB-404 Blending Stock

FB-405 Benzene Storage

FB-406 Toluene Storage

FB-407 Xylene Storage

FB-409 Gasoline Storage

FB-410 Neutral Oil Storage

FB-411 Phenol Product Storage

FB-412 Crude Cresylic Acid Storage

FB-413 O-Cresol Storage

FB-414 M, P Cresol Storage

FB-415 Xylenol Storage

3.0 CAPITAL COSTS

3.1 Equipment List - cont'd

CASE 4 - PROFITABLE JP-8

<u>TAG NO.</u>	<u>DESCRIPTION</u>
<u>AREA 400</u>	<u>STORAGE AREA</u>
GA-401A/S	Tar/Tar Oil Feed Pump
GA-402A/S	Crude Phenol Feed Pump
GA-403A/S	Fuel Oil Transfer Pump
GA-404A/S	Naphtha Transfer Pump
GA-405A/S	Crude Naphtha Transfer Pump
GA-406A/S	Blending Stock Pump
GA-407A/S	Benzene Transfer Pump
GA-408A/S	Toluene Transfer Pump
GA-411A/S	Gasoline Transfer Pump
GA-412A/S	Neutral Oil Transfer Pump
GA-413A/S	Crude Cresylic Acid Transfer Pump
GA-414A/S	O-Cresol Transfer Pump
GA-415A/S	M, P. Cresol Transfer Pump
GA-416A/S	Xylenol Transfer Pump

PA-401 Gasoline Blending Package

AREA 500 - CATALYST HANDLING

See Case 3 Area 500

AREA 600 - NAPHTHA DISTILL. AND HDT

See Case 7 Area 600

AREA 700 - AROMATICS RECOVERY

See Case 7 Area 700

AREA 800 - PHENOL EXTRACTION

See Case 7 Area 800

AREA 900 - CRESYLIC ACID DISTILLATION

See Case 7 Area 900

3.2 Cost Estimate

3.2.1 Basis of Estimate

The estimate for this case is a factored type estimate using the T.I.C. values developed for the various cases referenced in this project.

The total investment costs are scaled to the capacity requirement of this case use a 0.6 exponent.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs for Areas 100 thru 700 and 30% for Areas 800 & 900.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2 Estimate Summary

(Thousands of \$)

Case 4

Area 100 Hydrotreater	\$20,702
Area 200 Hydrocracker	10,012
Area 300 PSA & Recompression	8,400
Area 400 OSBL	9,421
Area 500 Catalyst Handling	1,285
Area 600 Naph. Dist & HDT	4,615
Area 700 ARU	7,887
Area 800 Phenol Ext.	12,276
Area 900 Cresylic Acid Dist.	4,832
Subtotal	\$79,430
Area 700 ARU Solvent Inventory	100
Total	\$79,530

3.2.3 Estimate Breakdown (Area 100) All Values in Thousands \$

This unit has the same capacity as the 100 Area of Case 3.
Therefore, T.I.C. = \$20,702.

Area 200

This unit has a 100% capacity of the 200 Area of Case 3.
Therefore, T.I.C. = \$10,012.

3.2.3 Estimate Breakdown - Cont'd

Area 300

This unit has a 102% capacity of the 300 Area of Case 3.
Therefore T.I.C. = $(1.02)^{0.6} (8300) = \$8400$.

Area 400

Tar Oil Stream Storage 100% Case 1
TIC = \$5,110

Phenol Stream Storage = 100% Case 7
TIC = \$3,016

Naphtha Stream Storage 100% Case 7
TIC = \$3058

Subtotal \$11,184

Less Duplicate Pipe & Rack -1763
Total = \$9,421

Area 500

The capacity of this unit is identical to the 500 Area of Case 3. Therefore T.I.C. = \$1,285

Area 600

This unit has a capacity of 100% of the 600 Area of Case 7.
Therefore T.I.C. = \$4,615

Area 700

This unit has a capacity of 100% of the 700 Area of Case 7.
Therefore T.I.C. = \$7,887

Area 800

This unit has a capacity of 100% of the 800 Area of Case 7.
Therefore T.I.C. = \$12,276

Area 900

This unit has a capacity of 100% of the 900 Area of Case 7.
Therefore = T.I.C. = \$4,832

4.0 OPERATING COSTS

4.1 Operating Labor

It is estimated that it will require men/shift to operate the plant broken down as follows:

Foreman	2	
Control Room	2	
HDT Operator	2	
HCR Operator	2	
PSA & relief man	1	
Naphtha Distil. & HDT	2	
ARU	2	
Phenol Ext.	1	
Cresylic Acid Dist.	<u>1</u>	
	15	Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 7 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	15 positions x 4 people/position -	60
Supervisor & Admin.		6
QC Technician		2
Maintenance		7
Other (Stores or Janitorial)		<u>1</u>
Total		76

4.2 Utilities

The following utilities have been estimated:

<u>Utility</u>	<u>Consumption</u>	<u>Cost</u>	<u>\$/SD</u>
#6 Fuel Oil	4361 BPSD	\$16/Bbl (a)	69776
SNG equivalent	3.73 MMSCFD	\$3.80/MM Btu (b)	13892
of Syn Gas & Purge Gas			
Cooling Water	6050 GPM	\$0.155/MGal (c)	1350
Power	6160 kW	\$0.04/kWh (c)	5914
Process Water	24 GPM	\$0.45/MGal (c)	15

(a) Cut of 1% sulfur \$6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

4.2 Utilities - cont'd

(b) Memo from D. Daley of Burns & Roe to L. Lorenzo of DOE dated Oct. 20, 1987, reference DPD-87-863.

(c) ANG utility cost information dated 5/87.

4.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

<u>Catalyst & Chem.</u>	<u>Use</u>	<u>Cost</u>	<u>\$/SD</u>
Nap. HDT Cat	0.021 #/Bbl	\$3.00/#	33
HDT Cat.	0.30 #/Bbl	\$3.00/#	2864
HCR Cat.	0.0095 #/Bbl	\$6.00/#	47
Inhibitors	50 PPM	\$10/Gal	52
ARU Solvent	24 #/D	\$2.10/#	50
H ₂ SO ₄	7100 #/D	\$0.04/#	285
			<u>\$3331</u>

4.4 Maintenance Supplies

Maintenance supplies for hydrotreating operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units (excluding the ARU solvent inventory). On this basis the maintenance supplies would be

$$0.00005 \times 79,430,000 = \$3971/\text{SD}$$

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the production of JP-8 and by-product chemicals are proposed to be located to the east of the Phenosolvan Unit and Water Treatment Area of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 340' x 575' will be surrounded by an access road and will be divided by three central east-west roads. Areas 100 & 500 will be located to the north and Areas 200 & 300 south of Area 100, Areas 800 & 900 next and then Areas 600 & 700.

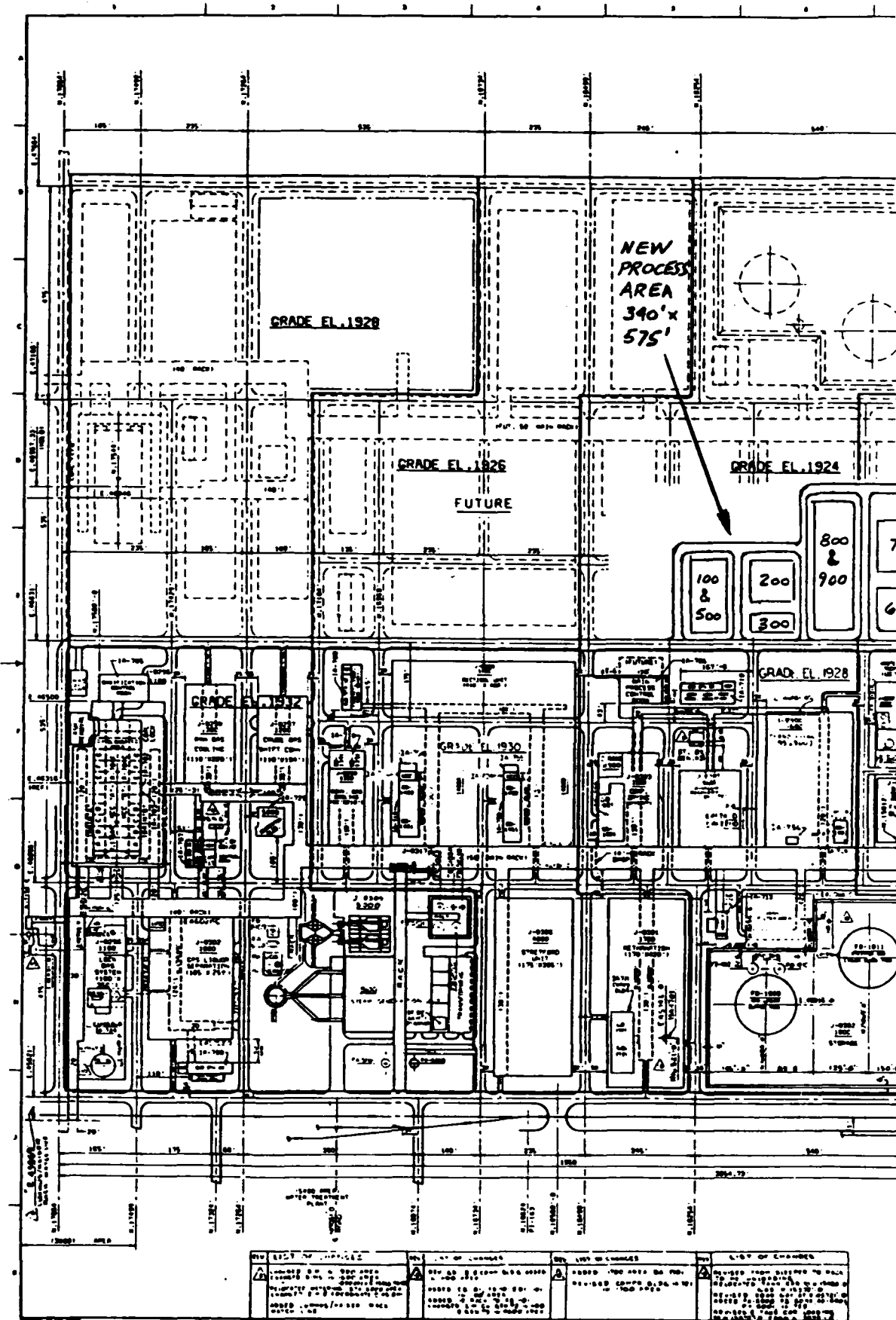
A diked storage tank area approx. 375' x 425' will be required for product and fuel oil storage and is proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

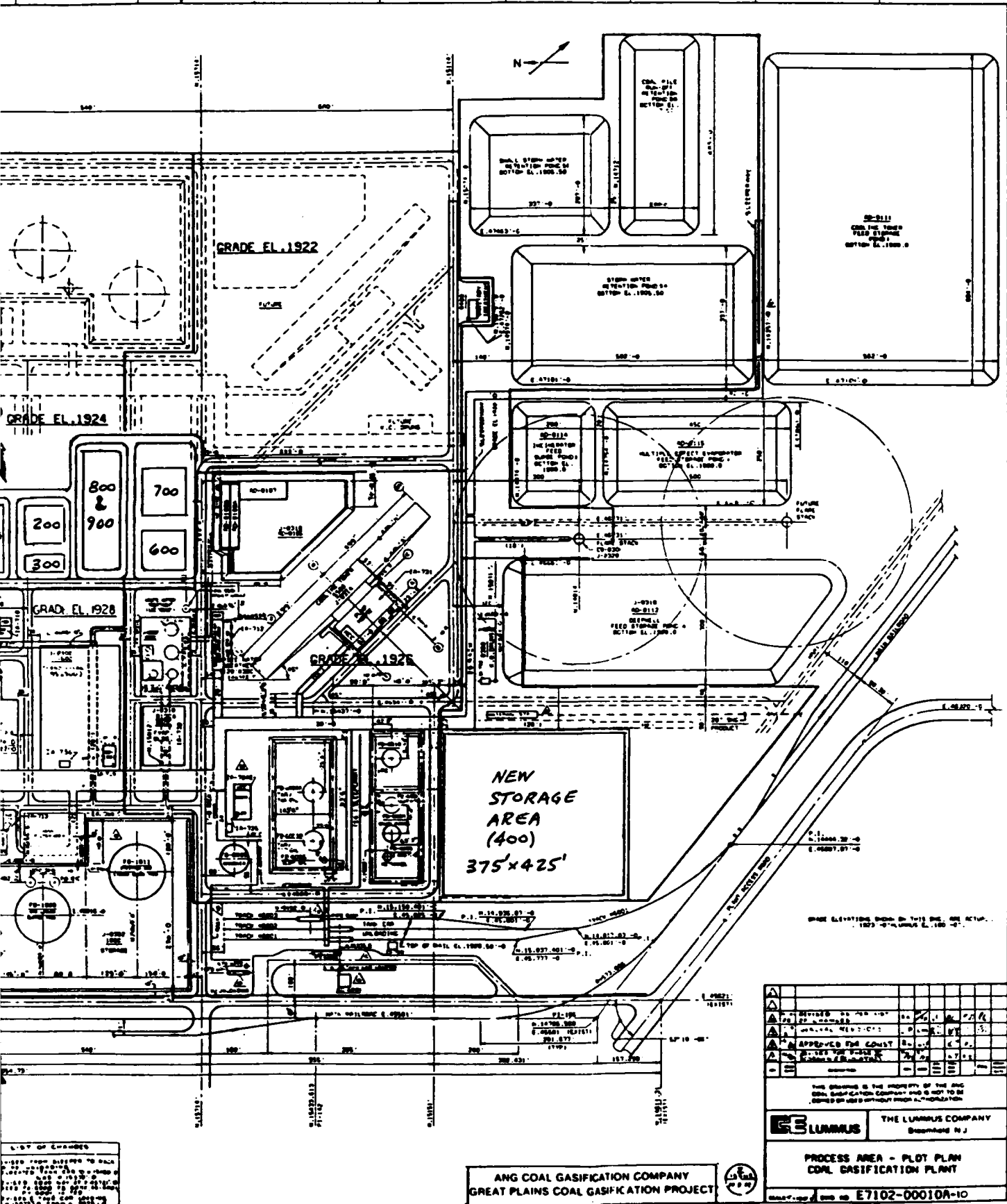
5.2 Unit Tie-Ins

Approximately 2500 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

A summary of the lines has not been prepared for this case but will be similar to a combination of Cases 3 and 7 with the utility lines of like services being combined.

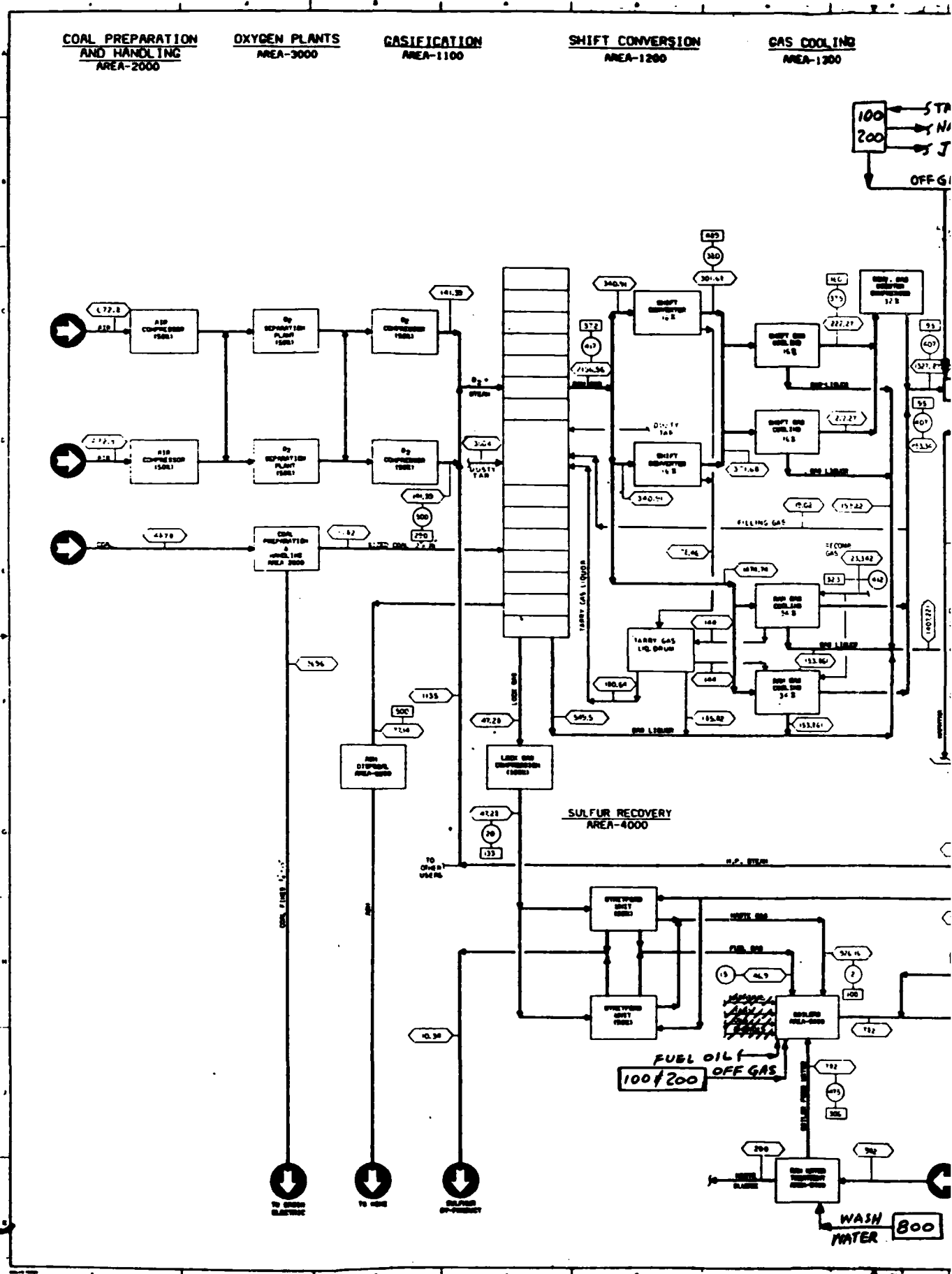




APPROVED FOR CONSTRUCTION	DATE	BY	FOR
APPROVED FOR DESIGN	DATE	BY	FOR
APPROVED FOR ERECTION	DATE	BY	FOR
APPROVED FOR COMPLETION	DATE	BY	FOR
THIS DRAWING IS THE PROPERTY OF THE ANG COAL GASIFICATION COMPANY AND IS NOT TO BE REPRODUCED OR USED IN ANY MANNER WITHOUT WRITTEN PERMISSION. THE LUMMUS COMPANY Birmingham, N.J. PROCESS AREA - PLOT PLAN COAL GASIFICATION PLANT Sheet No. E7102-00010A-10			

ANG COAL GASIFICATION COMPANY
GREAT PLAINS COAL GASIFICATION PROJECT

CASE 4



APPENDIX F

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 5
PROFITABLE JP-8X PRODUCTION
SUBTASKS 1.2

LCI PROJECT 5571
DATE - JAN. 30, 1988

F-1

W16248/bf

1.0 CASE DESCRIPTION

The design basis for the Maximum JP-8X case was received from Amoco and is presented on the following pages.

For various reasons, the most significant being the time restraint, this case was not developed into a conceptual process design and estimated by Lummus.

Also, of significance in the decision was the importance of developing Case 7 - Maximum Profit and that the Amoco LP did not choose vacuum distillation for the Profitable JP-8X Case 6.

FIGURE 1
BLOCK FLOW DIAGRAM
GREAT PLAINS CASE 5: MAXIMUM JP-8X PRODUCTION

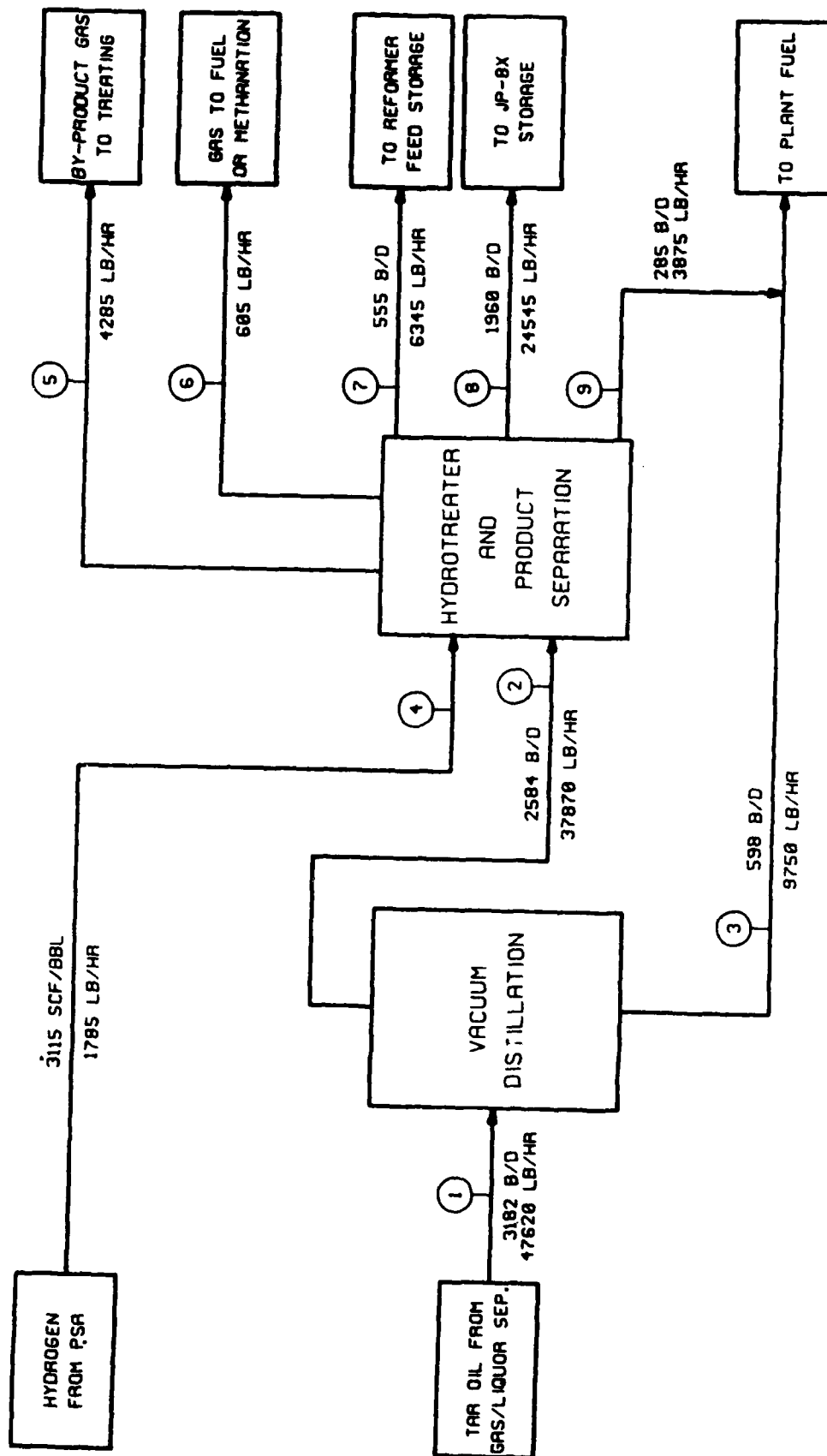


TABLE I
TAR OIL PROPERTIES
GREAT PLAINS CASE 5:
MAXIMUM JP-8X PRODUCTION

Flow Rate, B/D	3,182±500
API	6.6
Elemental Analysis, Wt%:	
C	82.1
H	8.6
N	0.9
S	0.4
O (Difference)	8.0
Solids, Wt%	0.8±0.6
Ash, Wt%	0.1±0.08
Ramscarbon Residue, Wt%	2.0
Water, Wt%	2.0±1.0
<u>Simulated Distillation*</u>	
Temperature @ IBP, °F	225
5%	320
10%	352
20%	405
30%	454
50%	542
70%	656
80%	734
90%	845
95%	918
FBP	1018

*D-2887 with aromatic standard

LPS/lkg/87863
9/9/87

TABLE II

STREAM FLOWS AND COMPOSITION
GREAT PLAINS CASE 5:
MAXIMUM JP-8X PRODUCTION

Stream No.	1	2	3	4	5	6	7	8	9
Description	Tar Oil	750-°F Tar Oil	750-°F Tar Oil	Hydrogen	By-product Gases	Hydrocarbon Gases	Reformer Feed	JP-8X	Hydro- treater 550-°F Product
Composition, Lb/Hr:									
H ₂	--	--	--	1785	--	55	--	--	--
C ₁	--	--	--	--	--	90	--	--	--
C ₂	--	--	--	--	--	145	--	--	--
C ₃	--	--	--	--	--	185	--	--	--
iC ₄	--	--	--	--	--	20	--	--	--
nC ₄	--	--	--	--	--	90	--	--	--
C ₅	--	--	--	--	--	20	--	--	--
H ₂ S	--	--	--	--	--	--	--	--	--
NH ₃	--	--	--	--	--	--	--	--	--
H ₂ O	950	950	--	--	160	--	--	--	--
Reformer Feed	--	--	--	--	340	--	--	--	--
JP-8X	--	--	--	--	3,785	--	6,345	24,545	--
Tar Oil	46,670	--	--	--	--	--	--	--	--
750-°F Tar Oil	--	36,920	--	--	--	--	--	--	--
Fuel Oil	--	--	9,750	--	--	--	--	--	--
Total Flow, Lb/Hr	47,620	37,870	9,750	1,785	4,285	605	6,345	24,545	3,875
B/D	3,182	2,584	598	--	--	60(FOE)	555	1,960	285

LPS/lkg/87663
9/9/87

TABLE III
PROPERTIES OF PRODUCTS FROM TAR OIL DISTILLATION
GREAT PLAINS CASE 5:
MAXIMUM JP-8K PRODUCTION

	<u>750-°F</u> <u>Tar Oil</u>	<u>750+°F</u> <u>Tar Oil</u>
Stream No.	2	3
Flow Rate, B/D	2,584	598
API	9.5	-4.7
Elemental Analysis, wt%:		
C	81.3	85.3
H	8.7	8.1
N	0.7	1.4
S	0.4	0.4
O (Difference)	8.9	4.8
Solids, wt%	0	3.9±2.9
Ash, wt%	0	0.5±0.4
Ramscarbon Residue, wt%	0	10
Water, wt%	2.5±1.0	0
<u>TBP Distillation, °F</u>		
IBP	272	557
5%	333	683
10%	363	714
20%	407	755
30%	445	785
50%	513	831
70%	587	859
80%	633	874
90%	690	890
95%	728	899
FBP	919	925

LPS/lkg/87863
9/9/87

TABLE IV
HYDROTREATER PRODUCT PROPERTIES
GREAT PLAINS CASE 5
MAXIMUM JP-8X PRODUCTION

	<u>Reformer Feed</u>	<u>JP-8X</u>	<u>550+°F</u>
Stream No.	7	8	9
Flow Rate, B/D	555	1,960	285
API	50	33	21
Composition, Vol%:			
Paraffins	14	9	20
Naphthenes	70	72	65
Aromatics	16	19	15
Octane: Research	72	--	--
Motor	68	--	--
RVP, psi	3	--	--
V/L = 20, °F	195	--	--
) ASTM D-86 IBP, °F	165	273	439
5%	180	343	490
10%	194	353	501
30%	232	380	529
50%	254	404	554
70%	273	433	617
90%	307	476	706
95%	325	495	732
FBP	392	596	767

LPS/lkg/87863
9/9/87

TABLE V
HYDROTREATER OPERATING CONDITIONS AND UTILITIES
GREAT PLAINS CASE 5
MAXIMUM JP-8X PRODUCTION

Operating Conditions

Reactor Type	Ebullated Bed
Catalyst	NiW
Feed Rate, B/D	3,200

Reactor Temperature, °F	700

Recycle Gas Rate, SCF/Bbl	6,000

Catalyst Replacement Rate, Lb/Bbl	0.18 (Approx. \$6/Lb)

Estimated Utilities

Fuel, MMBtu/Hr	0
Power, KW	2,600
Steam, MLb/Hr:	
600 psig	(650)
100 psig	250
Cooling Water, GPM	580
Process Water, GPM	150

LPS/lkg/87863
9/9/87

APPENDIX G

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

**CASE 6
PROFITABLE JP-8X PRODUCTION
SUBTASKS 1.2 & 1.3
PROCESS DESIGN AND COST ESTIMATE**

**LCI PROJECT 5571
DATE - JAN. 30, 1988**

W16246/bf

G-1

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4.0 OPERATING COSTS

- 4.1 Operating Labor
- 4.2 Utilities
- 4.3 Catalysts & Chemicals
- 4.4 Maintenance & Operating Supplies

5.0 PLOT PLAN & TIE INS

1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to produce JP-8X type aviation turbine fuel and chemical byproducts to maximize profit from Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- . A portion of the Tar Oil byproduct stream (40604 #/hr, 2713 BPSD) and the Neutral Oil stream produced in the Phenol Processing (4936 #/hr, 312 BPSD) is charged to the hydrotreater (Area 100).
- . The hydrotreater is a 3 stage expanded bed type process which removes 99% + of the sulfur, nitrogen, and oxygen compounds and begins the conversion of 550°F+ material. The hydrotreater adds a large quantity of hydrogen to the feed (3618 SCF/bbl) which results in a high heat of reaction. An expanded bed type reactor was chosen to both control and utilize the heat of reaction. Three stages were used to both control the temperature rise as well as to obtain the high efficiency associated with staging a back-mixed reactor.
- . The hydrotreater produces 6 streams:
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H₂ and CH₄.
 - Low pressure fuel gas (75 psig) which is sent to the main boiler in the SNG plant.
 - Unstabilized naphtha which is sent to the naphtha stabilizer. After stabilization, to control vapor pressure, the naphtha is sent to storage and gasoline blending.
 - JP-8X turbine fuel which is sent to storage.
 - 550°F+ unconverted bottoms product which is sent storage and fuel.
 - Wastewater containing, NH₄OH and NH₄HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H₂S and NH₃.

- Approximately 900 #/day of spent catalyst which is shipped to a catalyst reclaimer in the same drums that the catalyst is received in.
- . Hydrogen make-up for both the Hydrotreater (Area 100) and the Naphtha Hydrotreater (Area 600) is supplied from a PSA Hydrogen Unit (Area 300). High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining purge gas is available a low pressure (5 psig) which has a fuel value of about 565 BTU/ft³. This H₂, CO & CH₄ rich gas is recompressed into the methanation unit of the SNG plant.
- . The crude naphtha byproduct stream (8738#/hr, 725 BPSD) is charged to the distillation and hydrotreating unit (Area 600).
- . The distillation removes the material boiling below 160°F, which is sent to the SNG Plant fuel pool, and produces a bottoms product which is charged to the hydrotreater.
- . The fixed bed hydrotreater is a single bed reactor which removes 99% + of the sulfur, nitrogen, and oxygen compounds. Hydrogen is added to the feed at the rate of 430 SCF/bbl.
- . The naphtha hydrotreater produces 4 streams:
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H₂ and CH₄.
 - Naphtha which is stabilized to control vapor pressure, and the sent to the aromatics recovery unit (Area 700).
 - A low pressure off gas which is sent to the Stretford unit in the SNG plant.
 - Wastewater containing, NH₄OH and NH₄HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H₂S and NH₃.

The hydrotreated naphtha is charged to the extraction section of the Aromatics Recovery Unit (Area 700) where it is contacted with a solvent to extract the aromatic components from the stream. The raffinate is sent to storage and gasoline blending while the solvent is recovered from the aromatic extract. The aromatic extract is then sent to fractionation to produce the BTX products.

1.1 Overall Process Description - cont'd

- . Five streams are produced in the ARU plant.
 - A hydrocarbon gasoline blending stock which is sent to storage and gasoline blending.
 - A small process water stream which is sent to the waste treatment plant in the SNG Plant.
 - Three product streams Benzene, Toluene & Xylene which are sent to storage.
- . The crude phenol byproduct stream (14490 #/hr, 936 BPSD), is feed to the dual solvent phenol extraction unit (Area 800).
- . Distillation removes approx. 85% of the phenol which is further distilled to remove light ends and then reflashed over sulfuric acid producing a 99.8% pure product.
- . The remainder of the stream (a cresylic acid mixture) is flash distilled over a 3 wt.% concentrated sulfuric acid mixture to remove pyridine type substances.
- . The acid tar produced is water washed and mixed with light oil and sent to fuel.
- . The remaining cresol/xylenol mixture is double solvent extracted to remove neutral hydrocarbons. The resulting crude cresylic acid is dried and sent either to storage or distillation (Area 900).
- . Streams produced in the phenol extraction unit are:
 - Phenol product sent to storage
 - Crude Cresylic Acid sent to distillation (Area 900) or storage.
 - Wash Water sent to Water Treatment in the SNG Plant.
 - Waste Water sent to the Phenosolvan unit in the SNG Plant.
 - Neutral Oil sent to storage and fuel for the SNG Plant boilers.
- . The Crude Cresylic Acid is progressively distilled (Area 900) to separate the cresols and xylenols. No attempt has been made to remove the guaiacol from the product streams.

1.1 Overall Process Description - cont'd

- . Streams produced in the crude cresylic acid distillation unit are:
- o-Cresol product which is sent to storage.
 - m,p-Cresol product which is sent to storage.
 - Xylenol product which is sent to storage.
 - A heavy distillate which is combined with neutral oil in Area 800.
 - A crude phenol stream which is recycled to the Area 800.
 - A small water stream which is sent to Area 800 for tar acid washing.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas 160°F-distillate, and 550°F heavy oil produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

936 BPSD of Crude Phenol
725 BPSD of Crude Naphtha
2713 BPSD of Tar Oil
3312 BPSD of #6 Fuel Oil
8.72 MMSCFD equivalent SNG product loss due to the syn gas feed to the PSA unit.

Products

1938 BPSD of JP-8X turbine fuel
522 BPSD of 300°F - Naphtha for gasoline blending
317 BPSD of Phenol
56 BPSD of o-Cresol
131 BPSD of m,p-Cresol
75 BPSD of Xylenols
844 BPSD of 550°F+ for Fuel
202 BPSD of 160°F - Distillate for Fuel

1.2 Overall Material Balance - cont'd

Products

46 BPSD of Gasoline Blending Stock
315 BPSD of Benzene
112 BPSD of Toluene
15 BPSD of Xylene
5.60 MMSCFD equivalent SNG product credit due to HDT, & PSA
purge gas reinjection into SNG plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	3312 BPSD
SNG Equivalent of Syn Gas & Purge Gas	3.12 MM SCFD
Power	4810 kW
Cooling Water	5250 GPM (30°F rise)
Process Water	24 GPM

In addition the process utilizes steam as summarized below which was debited against boiler requirements.

HP Steam Import	58,200 #/H
MP Steam Import	9,300 #/H
LP Steam Export	7,050 #/H
Condensate Return	67,500 #/H

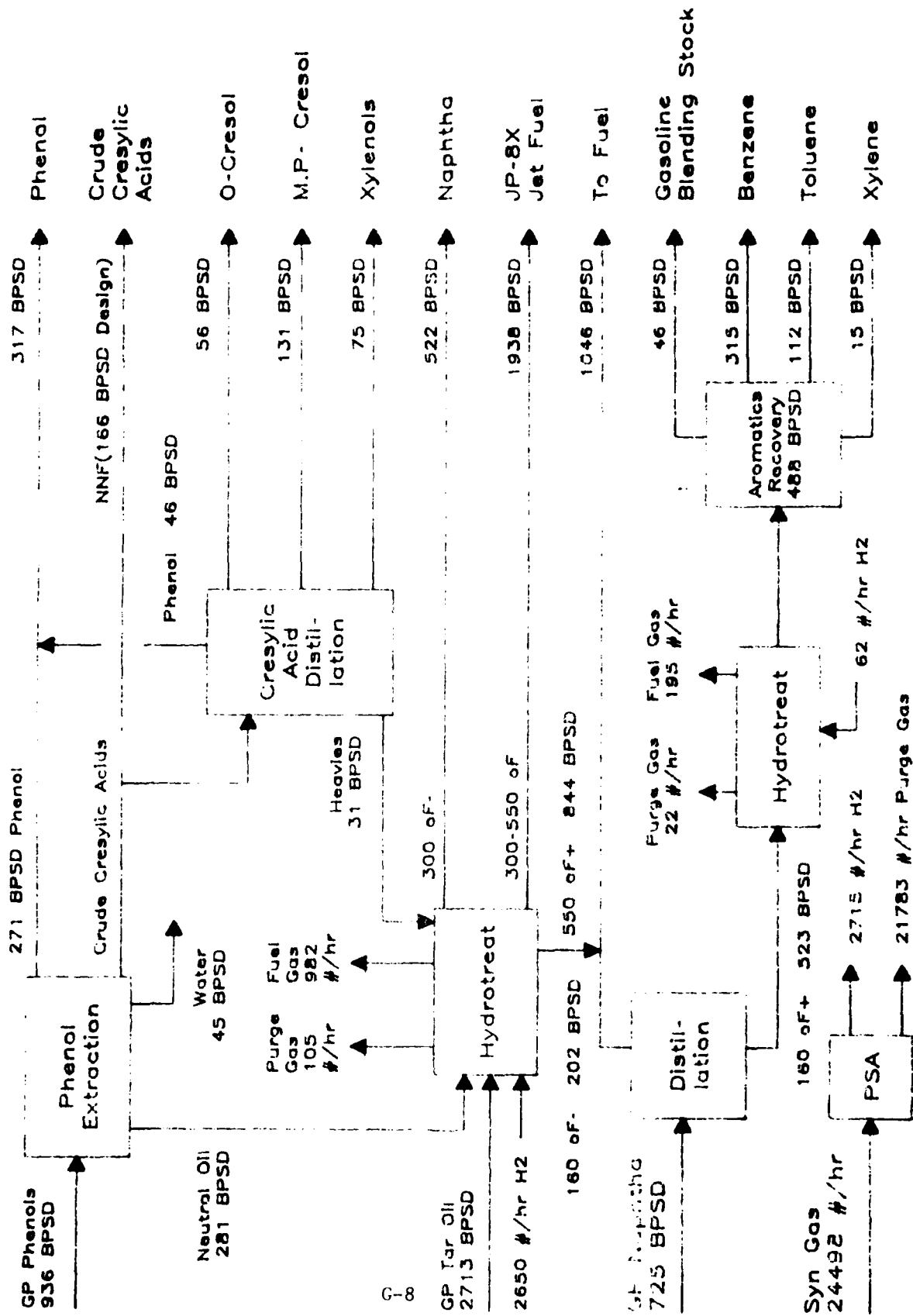


Table 1.1 Great Plains Case 6: Profitable JP8X Production

=====					
Tar Oil Feed====>	40604	#/hr	2713	BPSD	
Phenol Feed====>	14490	#/hr	936	BPSD	
Cr'd Naphtha Feed=>	8738	#/hr	725	BPSD	
Naphtha Product==>	5665	#/hr	522	BPSD	
JP8X Product====>	24318	#/hr	1938	BPSD	
Phenol Product===>	4925	#/hr	317	BPSD	
o-Cresol Prod====>	845	#/hr	56	BPSD	
m,p-Cresol Prod==>	1974	#/hr	131	BPSD	
Xylenols Prod====>	1070	#/hr	75	BPSD	
Gasoline Stock===>	481	#/hr	46	BPSD	
Benzene Prod====>	4060	#/hr	315	BPSD	
Toluene Prod====>	1425	#/hr	112	BPSD	
Xylene Prod=====>	188	#/hr	15	BPSD	
SNG Product Loss=>	5353	#/hr	3.1	MMSCFD	
Fuel Oil Makeup==>	45852	#/hr	3312	BPSD	

Expanded Bed Hydrotreater

=====					
Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD

Oil Feeds					
Tar Oil	89.16	1.0268	40604		2713
Neut. Oil	9.76	1.0860	4445		281
Heavies	1.08	1.0800	491		31

Tot. Oil	100.00		45540		3025
H2	5.82		2650	1314.5	

Total	105.82		48190		
Products					
Purge Gas	0.23		105	32.6	
Fuel Gas	2.16		982	48.4	
Naphtha	12.69	0.7600	5781		522
JP-8X	53.40	0.8608	24318		1938
S50 oF+	26.88	0.9950	12240		844
H2O in SW	8.96		4082	226.8	
H2S in SW	0.43		197	5.8	
NH3 in SW	1.06		485	28.5	

Total	105.82		48190		3304

Naphtha Stabilizer

=====					
Comp.	Wt %	Grav	#/hr	#Mole/hr	BPSD

HDT Nap	100.00	0.7600	5781		522

Stab Nap	98.00	0.7450	5665		522
Fuel Gas	2.00		116	2.7	

PSA Hydrogen Recovery Unit(86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	8.63	5.55	903.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	1564.3	460.9	36.6	401.4	7.8	4.7	2475.7
Prod. H2	1345.3	0.1	0.0	0.0	0.0	0.0	1345.4
Purge Gas	219.0	460.8	36.6	401.4	7.8	4.7	1130.3
#/hr							
Feed Gas	3154	12908	1612	6440	234	151	24498
Prod. H2	2712	3	0	0	0	0	2715
Purge Gas	442	12905	1612	6440	234	151	21783

Crude Naphtha
Distillation

	Wt %	Gravity	#/hr	BFSD
Feed Naphtha	100.00	0.8269	8738	725
Prod 160 oF-	24.77	0.7350	2164	202
Prod 160 oF+	75.23	0.8627	6574	523

Naphtha Hydrotrater

Component	Wt %	Grav	#/hr	#Mole/hr	BFSD
Feed 160 oF+	100.00	0.8627	6574		523
Feed Hydrogen	0.94		62	30.8	
Feed Total	100.94		6636		523
Products					
Purge Gas	0.33		22	6.8	
Fuel Gas	2.97		195	10.8	
HDT Naphtha	93.61	0.8650	6154		488
H2O in SW	1.96		129		
H2S in SW	1.76		116		
NH3 in SW	0.30		20		
Total Products	100.94		6636		488

Aromatics Recovery

Component	Wt %	Grav	#/hr	BPSD
Feed HDT Naphtha	100.00	0.8650	6154	488
Products				
Raffinate	7.82	0.7175	481	46
Benzene	65.97	0.8844	4060	315
Toluene	23.16	0.8718	1425	112
Xylene	3.05	0.8729	188	15
Total Products	100		6154	488

Phenol Extraction

Component	Wt %	#/hr	Grav	BPSD
Feeds				
Crude Phenol	100.00	14490	1.0621	936
Sulfuric Acid	1.97	285	1.8300	11
Total Feed	101.97	14775		947
Products				
Cr. Cresylic Acid	35.13	5090	1.0290	339
Phenol	29.09	4215	1.0661	271
Neutral Oil	30.68	4445	1.0860	281
Acidic Waste Water	7.07	1025	1.2558	56
Total Products	101.97	14775		947

Cresylic Acid Distillation

Component	Wt %	#/hr	Grav	BPSD
Cr. Cresylic Acid	100.00	5090	1.0290	339
Products				
Phenol	13.95	710	1.0661	46
o-Cresol	16.60	845	1.0350	56
m,p-Cresol	38.78	1974	1.0340	131
Xlenols	21.02	1070	0.9750	75
Heavies	9.65	491	1.0800	31
Total	100.00	5090		339

Fuel Gas Generated in Hydrotreating and Hydrocracking

Component	#/hr	#Mol/hr	MMBTU/hr
HDTR FG Produced	982	48.4	17.7
Stabilizer FG	116	2.7	2.1
Naphtha Hdtr FG	195	10.8	3.5
Total Fuel Gas	1098	51.1	19.8

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft3	MMBTU/hr
H2	442	219.0	324	26.9
CO	12905	460.8	321	56.1
CO2	1612	36.6	0	0.0
C1	6440	401.4	1010	153.7
C2	234	7.8	1769	5.2
N2+Ar	151	4.7	0	0.0
Total	21783	1130.3	565	241.8

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft3	BPSD
Tar Oil	-40604	17000	-690.3			-2713
Crude Phenol	-14490	13070	-189.4			-936
Crude Naphtha	-8738	18500	-161.7			-725
Fuel Gas	1098	18000	19.8	0.5	1020	
160 oF- distillate	2164	17400	37.7			202
550 oF+Hvy Oil	12240	18000	220.3			844
Import Steam	-60450	1000	-60.5			
Fuel Oil to Boiler	45779	18000	824.0			3307
Total	-63001		0.0	0.5		-21
Fuel Oil to Process Heaters	73	18000	1.3			5

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	8.72	59416
SNG Credit for PSA Purge gas	5.46	27127
SNG Credit for Hdtrs purge gas	0.14	947
Total SNG Production Loss	3.12	

2.0 PROCESS DESCRIPTION

2.1 Hydrotreater (Area 100)

For a description of the Hydrotreater process see Case 3 Section 2.1.

2.2 PSA Hydrogen Unit & Recompression (Area 300)

2.2.1 Hydrogen for both hydrotreaters will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig
Temp.	65 °F
Composition	mol%
H2	63.19
CO	18.61
CO2	1.48
CH4	16.21
C2H6	0.31
COS, H2S, CS2	< 0.01
N2 + Ar	0.19
H2O	< 0.01

The PSA unit selectively absorbs all components except H2 and produces a 99.99% vol. purity stream at about 345 psig and 80°F. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100°F
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

At the conditions given a 8 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

2.3 PSA Hydrogen Unit & Recompression (Area 300) - cont'd

2.3.1 Cont'd

The system uses 8 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continuously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-60301 presents a schematic of a Union Carbide Polybed PSA unit.

2.3.2 The purge gas is recompressed to 375 psia and sent to the methanation unit of the SNG plant.

2.4 Phenol Stream (Area 800 and 900)

For a description of the Phenol Extraction (Area 800) and Cresylic Acid Distillation Units see Case 7 Section 2.1.

2.5 Naphtha Stream (Area 600 and 700)

For a description of the Naphtha Distillation and Hydrotreating Unit (Area 600) and the Aromatics Recovery Unit (Area 700) see Case 7 Section 2.2.

PAGE 2-3

TYPICAL ARRANGEMENT

NUMBER OF ABSORBERS FOR THIS CASE = 8

ARRANGEMENT		SUBS ORBERS FOR THIS CASE = 8	
△			
△	1.2	ML	E.S.
REV	DESCRIPTION	DATE	BY
1	FOR SUBTASK 1.2	1.2	1.2
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AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.1 Equipment List

CASE 6 - PROFITABLE JP-8X

AREA 100 - HYDROTREATER

TAG. NO. DESCRIPTION

See Case 3 Area 100

AREA 300 - PSA HYDROGEN UNIT & RECOMPRESSION

FA-301 Purge Gas Surge Drum

GB-301 Purge Gas Compressor

PA-301 PSA Hydrogen Unit Package

AREA 400 - STORAGE AREA

FB-401 Jet Fuel Storage Tank

FB-402 Naphtha Storage Tank

FB-403 Fuel Oil Storage Tank

FB-404 Blending Stock

FB-405 Benzene Storage

FB-406 Toluene Storage

FB-407 Xylene Storage

FB-409 Gasoline Storage

FB-411 Phenol Product Storage

FB-412 Crude Cresylic Acid Storage

FB-413 O-Cresol Storage

FB-414 M, P Cresol Storage

FB-415 Xylenol Storage

3.0 CAPITAL COSTS

3.1 Equipment List - cont'd

CASE 6 - PROFITABLE JP-8X

<u>TAG NO.</u>	<u>DESCRIPTION</u>
<u>AREA 400</u>	<u>STORAGE AREA</u>
GA-401A/S	Tar/Tar Oil Feed Pump
GA-402A/S	Crude Phenol Feed Pump
GA-403A/S	Fuel Oil Transfer Pump
GA-404A/S	Naphtha Transfer Pump
GA-405A/S	Crude Naphtha Transfer Pump
GA-406A/S	Blending Stock Pump
GA-407A/S	Benzene Transfer Pump
GA-408A/S	Toluene Transfer Pump
GA-409A/S	Xylene Transfer Pump
GA-411A/S	Gasoline Transfer Pump
GA-413A/S	Crude Cresylic Acid Transfer Pump
GA-414A/S	O-Cresol Transfer Pump
GA-415A/S	M, P. Cresol Transfer Pump
GA-416A/S	Xylenol Transfer Pump

PA-401 Gasoline Blending Package

AREA 500 - CATALYST HANDLING

See Case 3 Area 500

AREA 600 - NAPHTHA DISTILL. AND HDT

See Case 7 Area 600

AREA 700 - AROMATICS RECOVERY

See Case 7 Area 700

AREA 800 - PHENOL EXTRACTION

See Case 7 Area 800

AREA 900 - CRESYLIC ACID DISTILLATION

See Case 7 Area 900

3.2 Cost Estimate

3.2.1 Basis of Estimate

The estimate for this case is a factored type estimate using the T.I.C. values developed for the various cases referenced in this project.

The total investment costs are scaled to the capacity requirement of this case use a 0.6 exponent.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs for Areas 100 thru 700 and 30% for Areas 800 & 900.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2 Estimate Summary

(Thousands of \$)

	<u>Case 6</u>
Area 100 Hydrotreater	\$20,702
Area 300 PSA & Recompression	6,760
Area 400 OSBL	8,511
Area 500 Catalyst Handling	1,285
Area 600 Naph. Dist & HDT	4,615
Area 700 ARU	9,373
Area 800 Phenol Ext.	12,276
Area 900 Cresylic Acid Dist.	4,832
Subtotal	\$68,354
Area 700 ARU Solvent Inv.	100
Total	\$68,454

3.2.3 Estimate Breakdown (Area 100) All Values in Thousands

This unit has the same capacity as the 100 Area of Case 3.
Therefore, T.I.C. = \$20,702.

3.2.3 Estimate Breakdown - Cont'd

Area 300

This unit has a 71% capacity of the 300 Area of Case 3.
Therefore T.I.C. = $(0.71)^{0.6} (8300) = \$6760$.

Area 400

Tar Oil Stream Storage 75% Case 1
TIC = $(0.75)^{0.6} (5 \text{ m}) = \$4200$

Phenol Stream Storage = 100% Case 7
TIC = \$3,016

Naphtha Stream Storage 100% Case 7
TIC = \$3058

Subtotal \$10,274

Less Duplicate Pipe & Rack -1763
Total = \$8,511

Area 500

The capacity of this unit is identical to the 500 Area of Case 3. Therefore T.I.C. = \$1,285

Area 600

This unit has a capacity of 100% of the 600 Area of Case 7.
Therefore T.I.C. = \$4,615

Area 700

This unit has a capacity of 100% of the 700 Area of Case 7.
Therefore T.I.C. = \$9,373

Area 800

This unit has a capacity of 100% of the 800 Area of Case 7.
Therefore T.I.C. = \$12,276

Area 900

This unit has a capacity of 100% of the 900 Area of Case 7.
Therefore = T.I.C. = \$4,852

4.0 OPERATING COSTS

4.1 Operating Labor

It is estimated that it will require men/shift to operate the plant broken down as follows:

Foreman	2	
Control Room	2	
HDT Operator	2	
PSA & relief man	1	
Naphtha Distil. & HDT	2	
ARU	2	
Phenol Ext.	1	
Cresylic Acid Dist.	1	
	<u>13</u>	Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 7 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	13 positions x 4 people/position -	52
Supervisor & Admin.		6
QC Technician		2
Maintenance		7
Other (Stores or Janitorial)		<u>1</u>
Total		68

4.2 Utilities

The following utilities have been estimated:

<u>Utility</u>	<u>Consumption</u>	<u>Cost</u>	<u>\$/SD</u>
#6 Fuel Oil	3312 BPSD	\$16/Bbl (a)	52992
SNG equivalent	3.12 MMSCFD	\$3.80/MM Btu (b)	11619
of Syn Gas & Purge Gas			
Cooling Water	5250 GPM	\$0.155/MGal (c)	1172
Power	4810 kW	\$0.04/kWh (c)	4618
Process Water	24 GPM	\$0.45/MGal (c)	15

(a) Cut of 1% sulfur \$6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

4.2 Utilities - cont'd

(b) Memo from D. Daley of Burns & Roe to L. Lorenzo of DOE dated Oct. 20, 1987, reference DPD-87-863.

(c) ANG utility cost information dated 5/87.

4.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

<u>Catalyst & Chem.</u>	<u>Use</u>	<u>Cost</u>	<u>\$/SD</u>
Nap. HDT Cal	0.021 #/Bbl	\$3.00/#	33
HDT Cat.	0.30 #/Bbl	\$3.00/#	2720
Inhibitors	50 PPM	\$10/Gal	52
ARU Solvent	18 #/D	\$2.10/#	50
H ₂ SO ₄	7100	\$0.04/#	285
			<u>\$3140</u>

4.4 Maintenance Supplies

Maintenance supplies for hydrotreating operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units. On this basis the maintenance supplies would be

$$0.00005 \times 68,354,000 = \$3418/SD$$

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the production of JP-8X and by-product chemicals are proposed to be located to the east of the Phenosolvan Unit and Water Treatment Area of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 340' x 390' will be surrounded by an access road and will be divided by two central east-west roads. Areas 100, 300 & 500 will be located to the north, Areas 800 & 900 next and then Areas 600 & 700.

A diked storage tank area approx. 375' x 325' will be required for product and fuel oil storage and is proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

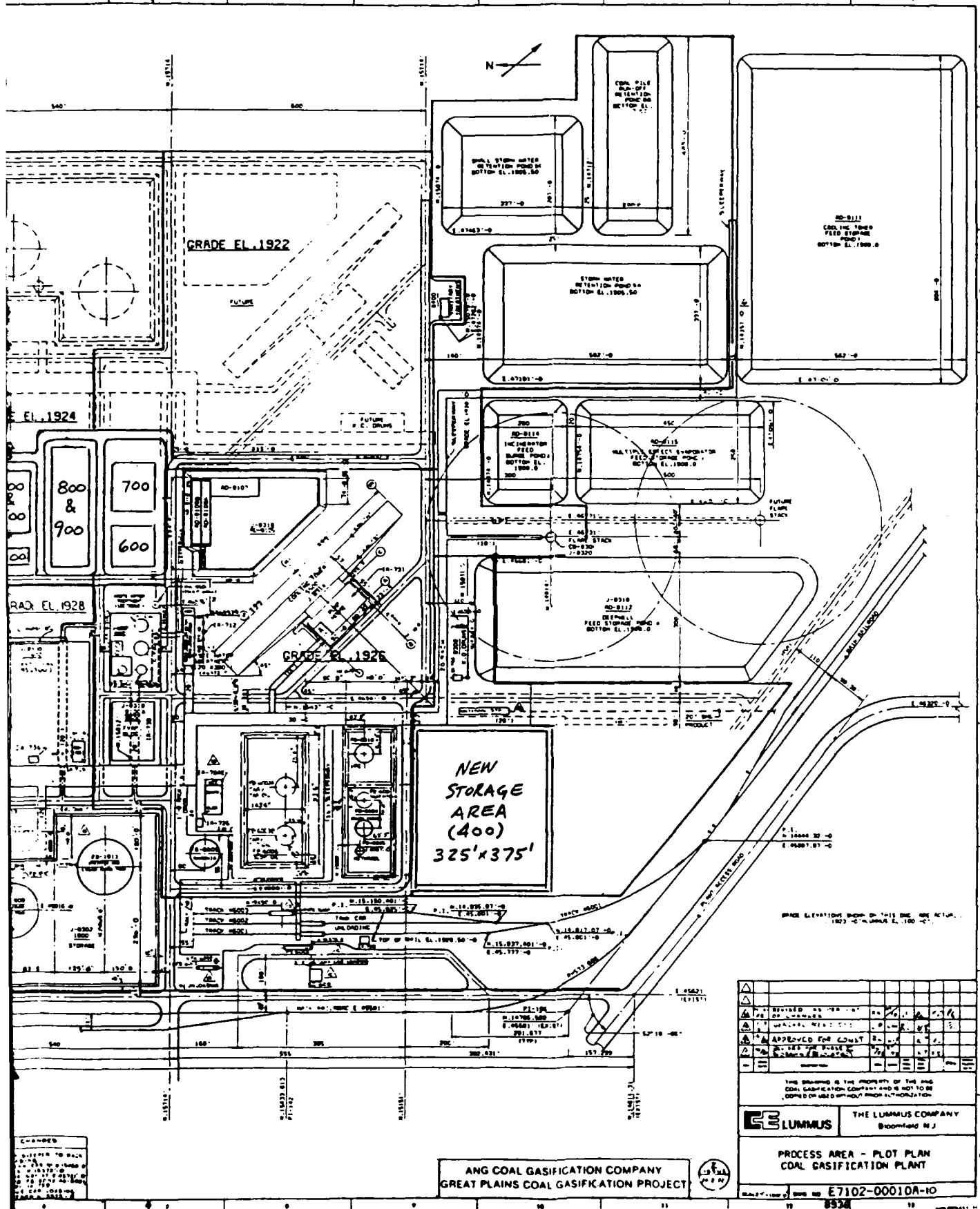
5.2 Unit Tie-Ins

Approximately 2500 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

A summary of the lines has not been prepared for this case but will be similar to a combination of Cases 3 and 7 with the utility lines of like services being combined.





CASE 6

TAR
NAP
JA
GA

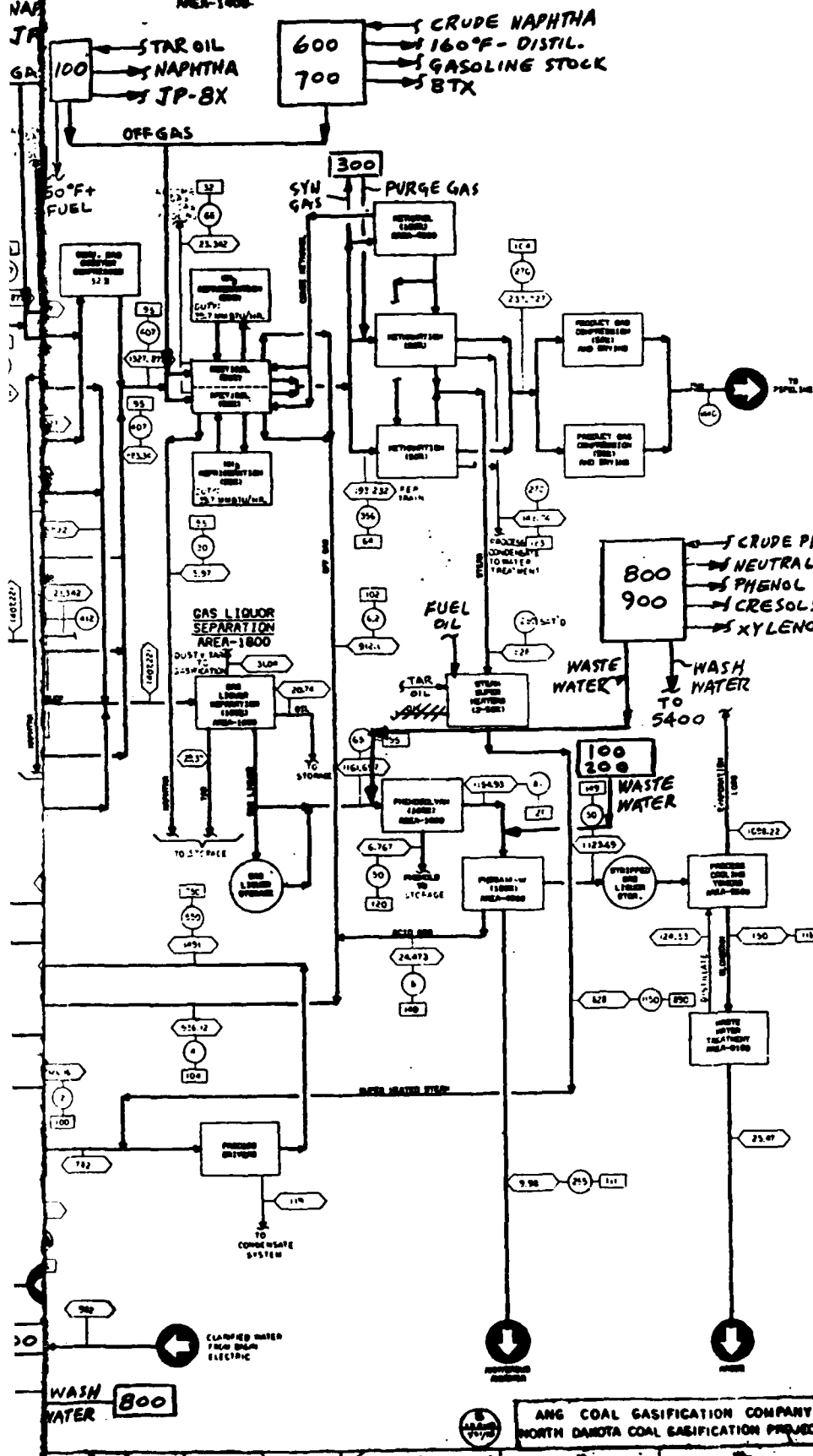
**ACID GAS
REMOVAL**
AREA-1000

METHANATION
AREA-1700

GAS COMPRESSION
AREA-1800

NOTES:

1. ALL FLOWRATES IN 10³ LB/HR.
2. TEMPERATURE IN °F
3. PRESSURE IN LB/IN² GAUGE



APPENDIX H

AMOCO/DOE

GREAT PLAINS GASIFICATION PLANT

JET FUEL FROM COAL DERIVED LIQUIDS

CASE 7
MAXIMUM PROFIT
SUBTASK 1.2
PROCESS DESIGN AND COST ESTIMATE

LCI PROJECT 5571
DATE JAN 30, 1988

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1.0 CASE DESCRIPTION

1.1 Overall Process Description

The purpose of this case is to maximize the profit from the processing of Great Plains liquid by products. Figure 1 presents a block diagram for the process and referring to Figure 1 the flow is as follows:

- . Two by product streams are processed.
 - Phenol 14490 #/hr (936 BPSD)
 - Naphtha 8738 #/hr (725 BPSD)
- . The crude naphtha byproduct stream (8738#/hr, 725 BPSD) is charged to the distillation and hydrotreating unit (Area 600).
- . The distillation removes the material boiling below 160°F, which is sent to the SNG Plant fuel pool, and produces a bottoms product which is charged to the hydrotreater.
- . The fixed bed hydrotreater is a single bed reactor which removes 99% + of the sulfur, nitrogen, and oxygen compounds. Hydrogen is added to the feed at the rate of 430 SCF/bbl.
- . The naphtha hydrotreater produces 4 streams:
 - High pressure purge gas (approximately 90% hydrogen) which is sent to the Rectisol Unit in the SNG plant for recovery of the H₂ and CH₄.
 - Naphtha which is stabilized to control vapor pressure, and then sent to the aromatics recovery unit (Area 700).
 - A low pressure off gas which is sent to the Stretford unit in the SNG plant.
 - Wastewater containing, NH₄OH and NH₄HS which is sent to the PHOSAM unit in the SNG plant for recovery of the H₂S and NH₃.
- . Hydrogen make-up for the Hydrotreater is supplied from a PSA Hydrogen Unit. High pressure (355 psig) synthesis gas from the Rectisol Unit (which contains about 63% hydrogen) is charged to the PSA unit which recovers 86% of the contained hydrogen as a high pressure 99.99% purity hydrogen gas product. The remaining gas is available at low pressure (5 psig) and has a fuel value of about 565 BTU/ft³. This gas is sent to the main boilers in the SNG plant.

- . The hydrotreated naphtha is charged to the extraction section of the Aromatics Recovery Unit (Area 700) where it is contacted with a solvent to extract the aromatic components from the stream. The raffinate is sent to storage and gasoline blending while the solvent is recovered from the aromatic extract. The aromatic extract is then sent to fractionation to produce the BTX products.
- . Five streams are produced in the ARU plant.
 - A hydrocarbon gasoline blending stock which is sent to storage and gasoline blending.
 - A small process water stream which is sent to the waste treatment plant in the SNG Plant.
 - Three product streams Benzene, Toluene & Xylene which are sent to storage.
- . The crude phenol byproduct stream (14490 #/hr, 936 BPSD), is feed to the dual solvent phenol extraction unit (Area 800).
- . Distillation removes approx. 85% of the phenol which is further distilled to remove light ends and then reflashed over sulfuric acid producing a 99.8% pure product.
- . The remainder of the stream (a cresylic acid mixture) is flash distilled over a 3 wt.% concentrated sulfuric acid mixture to remove pyridine type substances.
- . The acid tar produced is water washed and mixed with light oil and sent to fuel.
- . The remaining cresol/xlenol mixture is double solvent extracted to remove neutral hydrocarbons. The resulting crude cresylic acid is dried and sent either to storage or distillation (Area 900).
- . Streams produced in the phenol extraction unit are:
 - Phenol product sent to storage
 - Crude Cresylic Acid sent to distillation (Area 900) or storage.
 - Wash Water sent to Water Treatment in the SNG Plant.

1.1 Overall Process Description - cont'd

- Waste Water sent to the Phenosolvan unit in the SNG Plant.
- Neutral Oil sent to storage and fuel for the SNG Plant boilers.
- The Crude Cresylic Acid is progressively distilled (Area 900) to separate the cresols and xylenols. No attempt has been made to remove the guaiacol from the product streams.
- Streams produced in the crude cresylic acid distillation unit are:
 - o-Cresol product which is sent to storage.
 - m,p-Cresol product which is sent to storage.
 - Xylenol product which is sent to storage.
 - A heavy distillate which is combined with neutral oil in Area 800.
 - A crude phenol stream which is recycled to the Area 800.
 - A small water stream which is sent to Area 800 for tar acid washing.

1.2 Overall Material Balance

The overall material balance is presented in Table 1.1 which presents overall material balances for the major process units. Detailed material balances for the Phenol Extraction & Cresylic Acid Distillation can be found in Appendix A and for the Naphtha Distillation & HDT unit Appendix B. The balance was computed on the basis that the fuel value of the feed will be replaced by fuel gas, neutral oil and 160°F minus distillate produced in the process and the difference made up by the purchase of #6 Fuel Oil.

The overall balance is as follows:

Feeds

725	BPSD of Naphtha Feed
936	BPSD Phenol Feed
	BPSD of #6 Fuel Oil
0.20	MMSCFD equivalent SNG product loss due to the syn gas feed to the PSA unit.

1.2 Overall Material Balance - cont'd

Products

317 BPSD of Phenol
56 BPSD of o-Cresol
131 BPSD of m,p-Cresol
75 BPSD of Xylenols
312 BPSD of Neutral Oil for Fuel
202 BPSD of 160°F - Distillate for Fuel
46 BPSD of Gasoline Blending Stock
315 BPSD of Benzene
112 BPSD of Toluene
15 BPSD of Xylene

0.03 MMSCFD Equivalent SNG product credit due to HDT purge gas return to SNG plant.

1.3 Overall Utility Balance

The overall utility consumption of the complex is as follows:

#6 Fuel Oil	1193 BPSD
SNG Equivalent of Syn Gas	0.17 MM SCFD
Power	525 kW
Cooling Water	4000 GPM (30°F rise)
Process Water	3 GPM

In addition the process imports/exports steam and returns condensate as follows:

HP Steam	58,200 #/HR Import
MP Steam	15,900 #/HR Import
LP Steam	6,900 #/HR Export
Condensate return	74,100 #/HR

Figure 1 Case 7: Maximum Profit

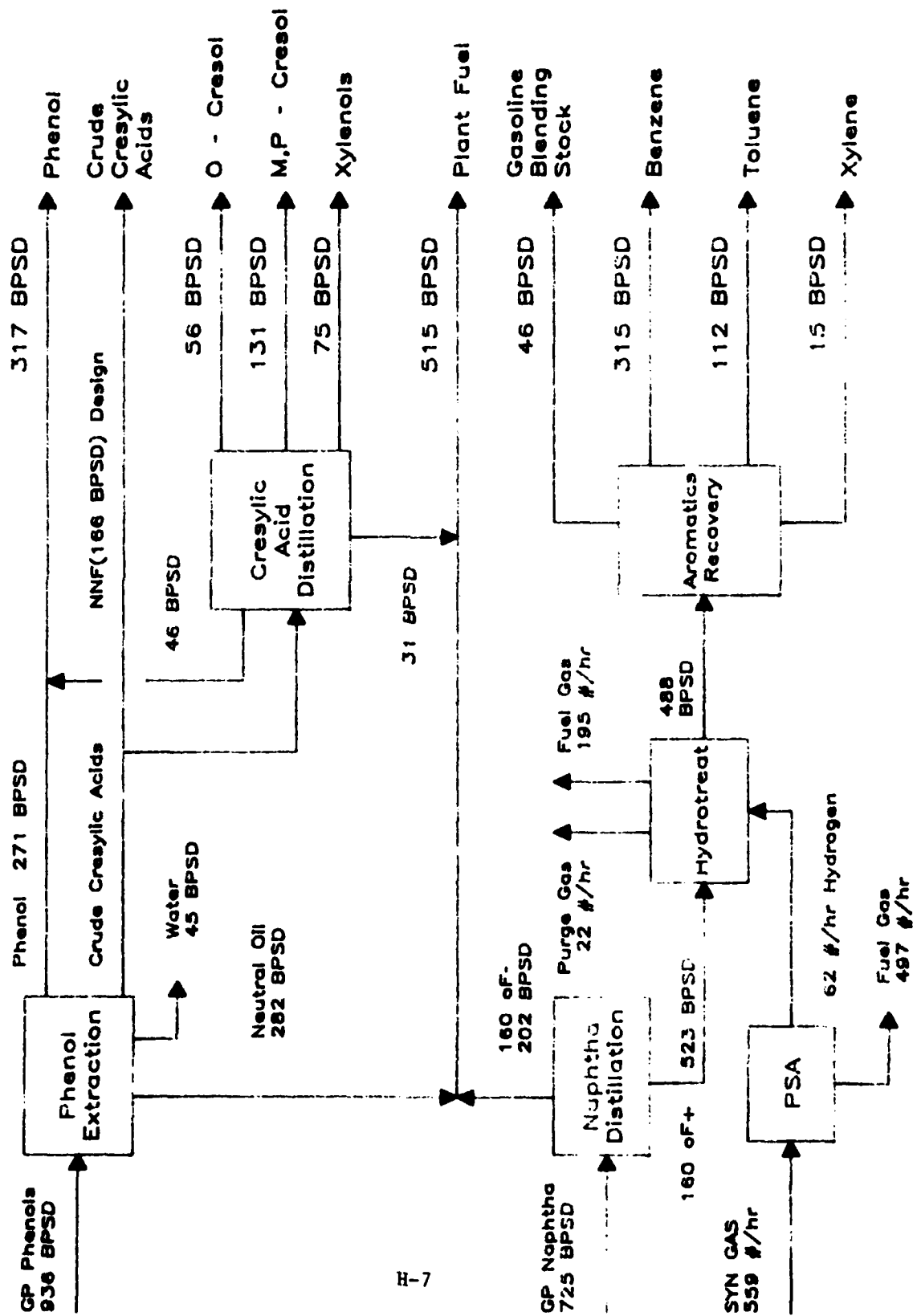


Table 1.1 Great Plains Case 7:Maximum Profit

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Phenol Feed====>	14490	#/hr	936	BPSD
Ord Naphtha Feed=>	8738	#/hr	725	BPSD
Phenol Product===>	4925	#/hr	317	BPSD
o-Cresol Prod====>	845	#/hr	56	BPSD
m,p-Cresol Prod==>	1974	#/hr	131	BPSD
Xylenols Prod====>	1070	#/hr	75	BPSD
Gasoline Stock===>	481	#/hr	46	BPSD
Benzene Prod====>	4060	#/hr	315	BPSD
Toluene Prod====>	1425	#/hr	112	BPSD
Xylene Prod=====>	188	#/hr	15	BPSD
SN6 Product Loss=>	300	#/hr	0.17	MMSCFD
Fuel Oil Makeup==>	16517	#/hr	1193	BPSD

Crude Naphtha Distillation =====	Wt %	Gravity	#/hr	BPSD
Feed Naphtha	100.00	0.8269	8738	725
Prod 160 oF-	24.77	0.7350	2164	202
Prod 160 oF+	75.23	0.8627	6574	523

Naphtha Hydrotrater

Component	Wt %	Grav	#/hr	#Mole/hr	BPSD
Feed 160 oF+	100.00	0.8627	6574		523
Feed Hydrogen	0.94		62	30.8	
Feed Total	100.94		6636		523
Products					
Purge Gas	0.33		22	6.8	
Fuel Gas	2.97		195	10.8	
HDT Naphtha	93.61	0.8650	6154		488
H2O in SW	1.96		129		
H2S in SW	1.76		116		
NH3 in SW	0.30		20		
Total Products	100.94		6636		488

PSA Hydrogen Recovery Unit(86% Recovery)

Component	H2	CO	CO2	CH4	C2H6	N2+Ar	Total
Mol %							
Feed Gas	116.28	34.26	2.72	29.84	0.58	0.35	184.03
Prod. H2	100.00	0.01					100.01
Purge Gas	16.28	34.25	2.72	29.84	0.58	0.35	84.02
Wt %							
Feed Gas	116.28	475.95	59.44	237.46	8.63	5.55	903.31
Prod. H2	100.00	0.12	0.00	0.00	0.00	0.00	100.12
Purge Gas	16.28	475.83	59.44	237.46	8.63	5.55	803.19
#Mol/hr							
Feed Gas	35.8	10.5	0.8	9.2	0.2	0.1	56.6
Prod. H2	30.8	0.0	0.0	0.0	0.0	0.0	30.8
Purge Gas	5.0	10.5	0.8	9.2	0.2	0.1	25.9
#/hr							
Feed Gas	72	295	37	147	5	3	560
Prod. H2	62	0	0	0	0	0	62
Purge Gas	10	295	37	147	5	3	498

Aromatics Recovery

Component	Wt %	Grav	#/hr	BFSD
Feed HDT Naphtha	100.00	0.8650	6154	488
Products				
Raffinate	7.82	0.7175	481	46
Benzene	65.97	0.8840	4060	315
Toluene	23.15	0.8715	1425	112
Xylene	3.05	0.8729	188	15
Total Products	100		6154	488

Phenol Extraction

Component	Wt %	#/hr	Grav	BFSD
Feeds				
Crude Phenol	100.00	14490	1.0621	936
Sulfuric Acid	1.97	285	1.8300	11
Total Feed	101.97	14775		947
Products				
Cr. Cresylic Acid	35.13	5090	1.0290	339
Phenol	29.09	4215	1.0661	271
Neutral Oil	30.68	4445	1.0860	281
Acidic Waste Water	7.07	1025	1.2558	56
Total Products	101.97	14775		947

Cresylic Acid Distillation

Component	Wt %	#/hr	Grav	BPSD
Feed Cresylic Acid	100.00	5090	1.0290	339
Products				
Phenol	13.95	710	1.0661	46
o-Cresol	16.60	845	1.0350	56
m,p-Cresol	38.78	1974	1.0340	131
Xylenols	21.02	1070	0.9750	75
Heavies	9.65	491	1.0800	31
Total	100.00	5090		339

Fuel Gas Generated in Hydrotreating

Component	#/hr	#Mol/hr	MMBTU/hr
Naphtha Hdttr FG	195	10.8	3.7

Purge Gas Generated in PSA Hydrogen Unit

Component	#/hr	#Mol/hr	BTU/ft ³	MMBTU/hr
H ₂	10	5.0	324	0.6
CO	295	10.5	321	1.3
CO ₂	37	0.8	0	0.0
C ₁	147	9.2	1010	3.5
C ₂	5	0.2	1769	0.1
N ₂ +Ar	3	0.1	0	0.0
Total	498	25.9	565	5.5

Net Changes in Boiler Fuel Fired

Fuel	#/hr	BTU/#	MMBTU/hr	MMSCFD	BTU/ft ³	BPSD
Crude Phenol	-14490	13070	-189.4			-936
Crude Naphtha	-8738	18500	-161.7			-725
PSA Purge Gas	498	11102	5.5	0.24	565	
Fuel Gas	195	19000	3.7	0.10		
160 oF- distillate	2164	17400	37.7			202
Neutral Oil	4936	15000	74.0			312
Import Steam	-67200	1000	-67.2			
Fuel Oil to Boiler	16517	18000	297.3			1193
Total	-66118		0.0	0.3		46

Net Changes in SNG Production	EQV SNG MMSCFD	PSA/Purge Gas #Mol/SD
SNG equivalent of Syn Gas to PSA	0.20	1359
SNG Credit for Hdttr purge gas	0.02	164
Total SNG Production Loss	0.17	

2.0 PROCESS DESCRIPTION

2.1 Phenol Stream

2.1.1 Phenol Extraction

Data for the design of the Phenol Extraction Unit (Area 800) were provided to Lummus by ANG. The basic processing step used in this unit is a dual solvent extraction to recover the phenol product. Referring to drawing D5571-70801A and B and the material balance (Appendix A) the flow is as follows:

- . Crude Phenol from the Great Plains Plant is charged to the Crude Phenol Column DA-801 from Surge Drum FA-801 through Feed Pump GA-801. Recycled phenol streams from Cresylic Acid Distillation (Area 900) and the Phenol Column DA-804 overheads are also charged to column DA-801. The bottoms from DA-801 is pumped by GA-802 to Acid Flash Column DA-802. The overhead from DA-802 contains light ends and phenol. Non condensibles are relieved to flare. The overhead liquids are pumped by GA-803 and GA-804 to the light ends column DA-803. Water condensed in the overheads is separated and sent to the Phenosolvan Unit in the SNG Plant.
- . The Crude Phenol Column overhead enters the Light Ends Column DA-803. The overhead light ends product is sent to SNG Plant Fuel. Water condensed in the overhead drum is sent to the Phenosolvan Unit in the SNG Plant. The bottoms from the Light Ends Column are pumped to Phenol Column via Light Ends Bottoms Pump GA-809.
- . Bottoms from the Light Ends Column are distilled in the Phenol Column DA-804 to produce an overhead stream which is returned to the Crude Phenol Column, DA-801 a side draw stream of 99.8% pure phenol product, and a bottoms stream which is pumped to Acid Flash Column DA-802 via Phenol Column Bottoms Pump GA-813. A small amount of sulfuric acid is added to the tower in order to control product purity.
- . The Phenol Product is drawn off DA-804 to Phenol Draw-off Pot FA-804 and pumped by Phenol Drawoff Pump GA-814 through Phenol Product Cooler EA-810 to Phenol Product Day Tank FB-803. From the day tank, phenol product is pumped to storage via GA-815, Phenol Product Pump.

2.0 PROCESS DESCRIPTION

2.1.1 Phenol Extraction - cont'd

- . The bottom streams from DA-801, Crude Phenol Column, and DA-804 Phenol Column are sent to Acid Flash Column DA-802. This combined stream contains the cresylic acid, neutral oil and heavies. This material is flashed over sulfuric acid to remove pyridine type substances. The overhead product of the acid flash is pumped to Extraction Column DA-805. The bottoms product is an acid tar, and is water washed in FD-801 & FD-802, 1st and 2nd stage Water Wash Tanks, to remove acid materials, and then routed to the Great Plains Fuel pool together with other fuel streams recovered in the Phenol Extraction and Cresylic Acid Distillation Units, (Areas 800 and 900). The combined stream is called "neutral oil."
- . The acid flash overhead from DA-804 is extracted with hexane and methanol/water in Extractor Column DA-805. Hexane enters the extractor column at the bottom and preferentially absorbs the oil components. The hexane/oil mixture exits the top of DA-805.
- . Methanol/water solution enters the top of the Extractor Column. The methanol/water preferentially adsorbs the phenolic compounds. Methanol/water/phenolic mixture exits the bottom of DA-805.
- . The oil components are stripped from the hexane in the Hexane Column DA-806. The hexane is recycled to Extractor Column DA-804. The oil is pumped by Hexane Column Bottoms Pump GA-822 through Neutral Oil Cooler EA-815 to Neutral Oil Day Tank FB-802. From FB-802 Neutral Oil can be pumped to fuel or storage by GA-812 Neutral Oil Pump.
- . Make-up Hexane is added as needed from Hexane Storage Tank FB-804 by GA-824 Hexane Make-up Pump.

2.0 PROCESS DESCRIPTION

2.1.1 Phenol Extraction - cont'd

- . The phenolics are recovered from the methanol/water solution in Methanol Column DA-807. The methanol/water is condensed overhead and recycled to the extractor column by Methanol Column Reflux Pump GA-826. The phenolics are pumped to Drying Column DA-808 by Methanol Column Bottoms Pump GA-825 through Methanol Column Bottoms Cooler EA-818.
- . Drying Column DA-808 is reboiled to remove water carry-over from the phenolic product. The dry Crude Cresylic Acid leaves the bottom of DA-808. Product is pumped to either Cresylic Acid Distillation (Area 900) or through the Crude Cresylic Acid Cooler EA-821 to the Cresylic Acid Day Tank, FB-805 by Drying Column Bottoms Pump GA-828.
- . Crude Cresylic Acid from Cresylic Acid Day Tank is pumped to storage by Crude Cresylic Acid Pump GA-829.

2.1.2 Cresylic Acid Distillation

Data for the design of the Cresylic Acid Distillation Unit (Area 900) were provided to Lummus by ANG. The basic process used is a series of distillation columns to recover progressively higher boiling products. Referring to drawing D5571-70901 and the material balance (Appendix A) the flow is as follows:

- . Dry Crude Cresylic Acid from Phenol Extraction (Area 800) is charged to O-Cresol Column DA-901.
- . o-Cresol Reboiler EA-901 uses HP Steam to reboil the column. Liquid distillate product is returned to the Crude Phenol Column in Area 800. o-Cresol is a side draw product. Column DA-901 bottoms are feed to m,p-Cresol Column DA-902.
- . o-Cresol product is stripped in o-Cresol stripper, DA-902, o-Cresol product is pumped by o-Cresol Stripper Pump GA-902 through o-Cresol Product Cooler EA-903 to o-Cresol Day Tank FB-901, o-Cresol product is pumped to storage by o-Cresol Product Pump GA-910.

2.0 PROCESS DESCRIPTION

2.1.2 Phenol Extraction - cont'd

- o-Cresol column bottoms is charged to the m,p-Cresol Column.
- m,p-Cresol product is recovered overhead. m,p-Cresol is pumped from m,p-Cresol Reflux Drum FA-902 by m,p-Cresol Reflux Pump GA-906 through m,p-Cresol Product Cooler EA-908 to the day tank FB-902. m,p-Cresol product is pumped to storage by the product pump GA-911.
- m,p-Cresol column bottoms is feed to DA-905 xlenol column. The xlenol product is recovered overhead. Xlenol is pumped from Xlenol Reflux Drum FA-903 by Xylene Reflux Pump GA-908 through Xlenol Product Cooler EA-911 to Xlenol Day Tank FB-903. Xlenol product is pumped to storage by Xlenol Product Pump GA-912.
- Xlenol column bottoms contains the undesirable heavies. Xlenol column bottoms are pumped by Xlenol Bottoms Pump GA-907 through Cresylic Acid By-product Cooler EA-912 to SNG plant fuel via the Neutral Oil system in Area 800.

2.2 Naphtha Stream

2.2.1 Naphtha Distillation & Hydrotreating (Area 600)

Operating conditions for the naphtha distillation and hydrotreater were provided to Lummus by Amoco and these conditions are presented in Table 2.1. The basic processing steps selected were a distillation to produce a 160°F+ feed stock and a fixed bed hydrotreater. Referring to drawing D5571-70601 and the material balance (Appendix B) the flow is as follows:

- The crude naphtha is charged to the Naphtha Distillation Column DA-601 via Surge Drum FA-601 and Feed Pump GA-601.
- The column is reboiled with steam in EA-601 to produce a 160°F+ bottoms product.
- The 160°F- overheads are condensed in EA-602 and sent to fuel via GA-603.
- The 160°F+ Distillation Column bottoms is charged to the HDT surge drum FA-603 via GA-602.

2.0 PROCESS DESCRIPTION

2.2.1 Hydrotreater - cont'd

- . 160°F+ naphtha is charged into the hydrotreater from surge tank FA-603 by charge pumps GA-604 through Feed/Effluent Exchanger EA-603.
- . The charge oil is combined with feed hydrogen gas from heater EA-604 prior to entering the feed/effluent exchanger. The preheated mixture is then charged to the reactor DC-601.
- . The reactor DC-601A operates adiabatically with an average bed temperature of 450°F.
- . The effluent from DC-601 is cooled in EA-603 and flows through exchangers EA-605 and EA-606. Process water is injected prior to EA-606 to convert the H₂S and NH₃ in the gas to an aqueous NH₄OH/NH₄HS solution.
- . The cooled mixture then passes into the High Pressure/Low Temperature Separator FA-605 where hydrogen rich gas leaves overhead. A portion of this high pressure gas is purged to remove H₂S and light gases from the loop and sent to the Rectisol Unit 1400 in the SNG plant to recover the hydrogen in the purge gas. The remaining gas is recirculated to reactor DC-601.
- . The water phase from separator FA-606 goes to the PHOSAM Unit in the SNG plant to recover the H₂S and NH₃.
- . The hydrocarbon phase from separator FA-606 is preheated in exchanger EA-605 and charged to the HDT Naphtha Stabilizer DA-602.
- . The unstabilized naphtha is charged into DA-602 which is reboiled by MP Steam to stabilize the naphtha.
- . Offgas from the Naphtha Stabilizer is sent to the SNG plant for fuel.
- . The stabilized naphtha is cooled and sent to the aromatics recovery unit (Area 700).

2.0 PROCESS DESCRIPTION

2.2.1 Hydrotreater - cont'd

Table 2.1 Hydrotreater Conditions

Case 7 Maximum Profit	
Feed Stock	160°F+ Naphtha
Reactor Type	Fixed Bed
Number of Stages	1
LHSV Hr ⁻¹	1.0
Average Reactor Temperature	450°F
Reactor Pressure	500 psig H ₂ Partial Pressure
H2 Recycle Rate	2500 SCF/BBL
Catalyst	Ni-Mo
Catalyst Replacement	2 years @ \$3/#

2.2.2 Aromatics Recovery Unit (Area 700)

This unit is based on the Shell Sulfolane Process licensed by Universal Oil Products. Referring to Drawings D5571-70701A and B the flow is as follows:-

Stabilized Naphtha from Naphtha Hydrotreater (Area 600) is charged to the Extraction Column DA-701 through Feed Surge Drum FA-701 by Feed Charge Pump GA-701. Lean Solvent is charged to the top of column DA-701. As the feed flows through the column, aromatic components are selectively dissolved in the solvent. Raffinate with very low aromatics content is withdrawn from the top of DA-701.

Rich solvent leaves the bottom of the extractor. After heat exchange in Lean/Rich Solvent Exchanger EA-702, the rich solvent is charged to the top of DA-703, Stripper.

The raffinate stream from the Extractor Column DA-701 overheads is cooled in raffinate cooler EA-701 and then contacted with wash water in Water Wash Column DA-702. Water removes any dissolved solvent from the raffinate. Raffinate leaving DA-702 overhead is pumped to Gasoline Blending Stock Storage. The solvent rich water from DA-702 flows to DA-705, Water Stripper.

Solvent accumulates in the bottom of Water Stripper DA-705 and is pumped back to the Recovery Column by Water Stripper Bottoms Pump GA-710. The rich water is returned to the Recovery Column as stripping steam generated via the Water Stripper Reboiler EA-709 by exchange with the hot circulating lean solvent.

2.2.2 Aromatics Recovery Unit (Area 700) - cont'd

A solvent regeneration system is included to guard against excessive solvent degradation. In normal operation a slipstream of solvent is routed to the Solvent Regenerator DA-706. Degraded solvent is periodically withdrawn from the bottom of DA-706.

In the Stripper, non-aromatic hydrocarbons, which are more volatile, are stripped from the solvent, removed overhead, condensed and recycled to the Extractor Column DA-701 for recovery.

The stripper bottoms consists of aromatics in the solvent. This stream is pumped to the Recovery Column DA-704 by Stripper Bottoms Pump GA-704.

In the Recovery Column DA-704, the aromatics are stripped from the solvent. Lean solvent leaves the column bottom and is returned to Extraction Column DA-701 by GA-707 Lean Solvent Pump after heat Exchange in Water Stripper Reboiler EA-709 and Lean/Rich Solvent Exchanger EA-702.

The aromatic product recovered overhead from the Recovery Column is fractionated to recover benzene, toluene and xylene product streams.

The recovery column overhead is pumped by Recovery Column Overhead Pump GA-709 to Clay Tower Surge Tank, FB-703. From FB-703 the aromatic stream is pumped by Clay Tower Feed Pump GA-715 through Clay Tower Feed/Effluent Exchanger EA-712 and Clay Tower Feed Heater EA-713 and then into Clay Towers DA-707A/B. In the Clay Tower, trace amounts of unsaturates and residual non-hydrocarbon impurities are removed.

After heat exchange in the Clay Tower Feed/Effluent Exchanger, the extract flows to benzene column DA-708. Benzene product is withdrawn from a tray near the top of the tower. After cooling in Benzene Product Cooler EA-715, benzene flows to Benzene Day Tank FB-704. Product from FB-704 is pumped to product storage by Benzene Product Pump GA-719.

Any water that accumulates in Benzene Column Reflux Drum FA-708 is permed to Waste Treatment by Benzene Column Water Pump GA-718.

2.2.2 Aromatics Recovery Unit (Area 700) - cont'd

Benzene column bottoms are pumped by Benzene Column Bottoms Pump GA-716 to Toluene column DA-709. The Toluene Product leaves overhead. Toluene is pumped from Toluene Column Reflux Drum FA-709 by Toluene Column Reflux Pump GA-721 through Toluene Product cooler EA-720 to Toluene Day Tanks FB-706A/B. Toluene from FB-706A/B is pumped to storage by Toluene Product Pump GA-723.

Xylene is taken as bottoms product from Toluene Column DA-709. Xylene is pumped by Toluene Column Bottoms pump GA-720 through Xylene Product Cooler EA-718 to Xylene Day Tank FB-705. Xylene from FB-705 is pumped to storage by Xylene Product Pump GA-722.

2.2.3 PSA Hydrogen Unit (Area 300)

Hydrogen for the naphtha hydrotreater will be supplied by a PSA hydrogen unit. The feed gas will come from the Rectisol Unit in the SNG plant (Stream ID-GF 1401) which has the following properties:

Pressure	355 psig
Temp.	65 °F
Composition	mol%
H2	63.19
CO	18.61
CO2	1.48
CH4	16.21
C2H6	0.31
COS, H2S, CS2	< 0.01
N2 + Ar	0.19
H2O	< 0.01

The PSA unit selectively absorbs all components except H2 and produces a 99.99% vol. purity stream at about 345 psig and 80°F. The other components are available as a purge gas having the following properties.

Pressure	5 psig
Temperature	100°F
Composition	Mole %
H2	19.32
CO	40.76
CO2	3.24
CH4	35.51
C2H6	0.69
N2+Ar	0.41
Others	0.06

2.0 PROCESS DESCRIPTION

2.2.3 PSA Hydrogen Unit (Area 300) - cont'd

At the conditions given a 4 bed PSA unit will recover 86% of the hydrogen in the feed according to the manufacturer, Union Carbide EP&P Division.

The system uses 4 absorption vessels which are sequenced through adsorption, depressurization, purging, and repressurization steps. The process continuously produces product and purge gas. It is purchased as a skid mounted unit and the control of the unit is fully automated. Drawing 5571-70301 presents a schematic of a Union Carbide Polybed PSA unit.

PAGE 2-10

TYPICAL ARRANGEMENT
NUMBER OF ABSORBERS FOR THIS CASE = 4

[illegible]

1090Z-1ZSSQ 'ON 'SHO

FA-601
CRUDE NAPHTHA
FEED SURGE DRUM

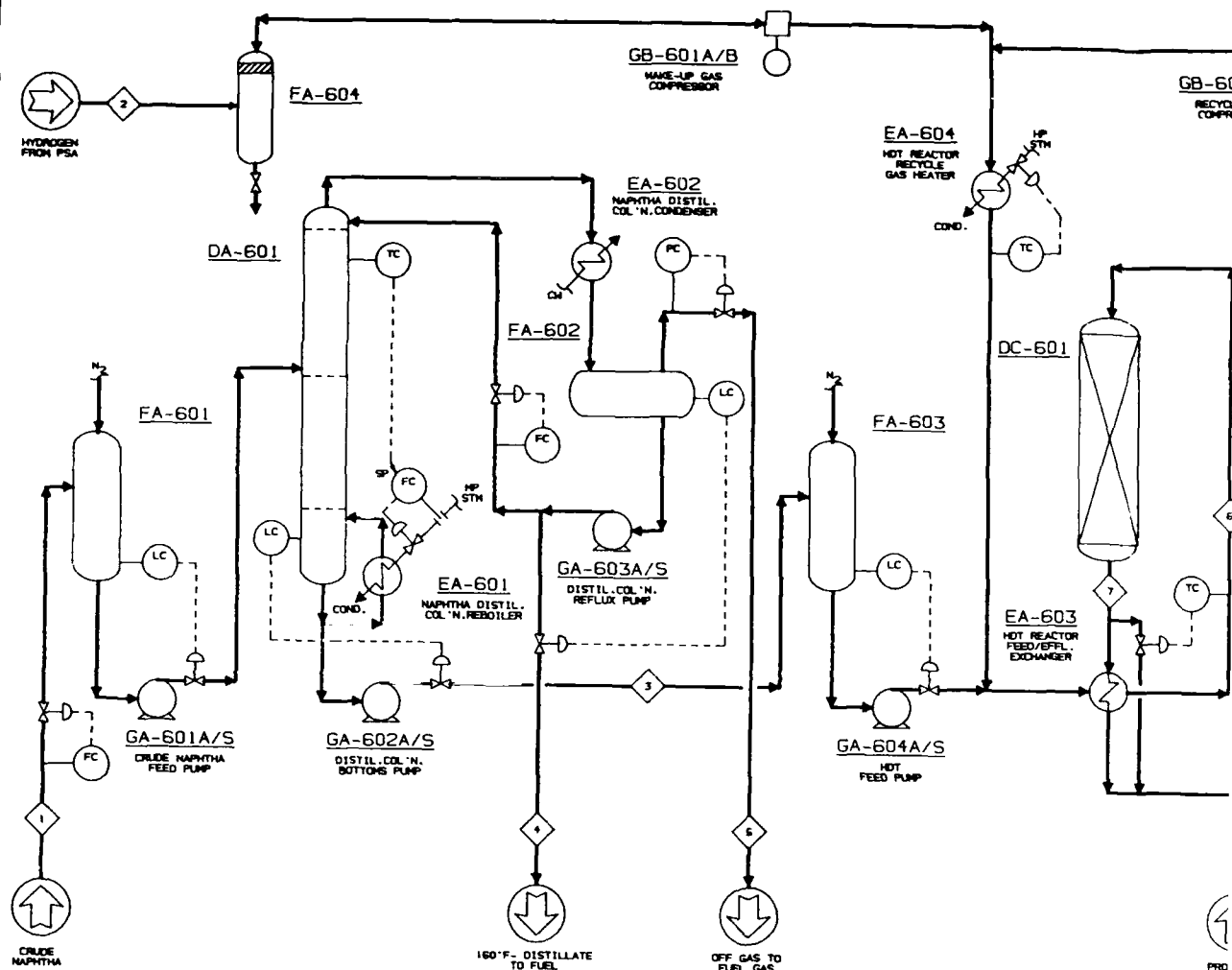
FA-604
MAKE-UP GAS
KO DRUM

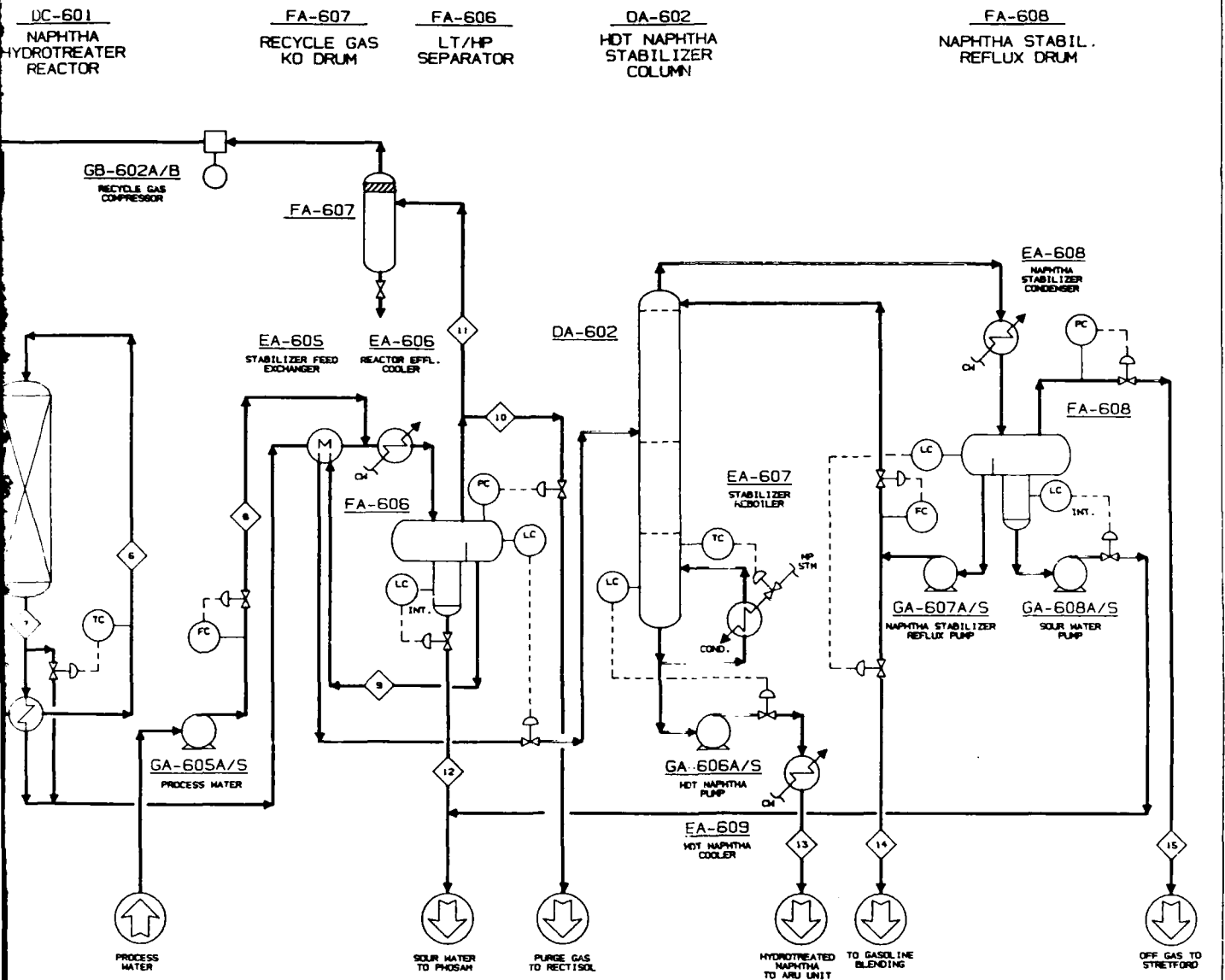
DA-601
NAPHTHA
DISTILLATION
COLUMN

FA-602
DIST. COL'N.
REFLUX DRUM

FA-603
HOT FEED
SURGE DRUM

DC-601
NAPHTHA
HYDROTREATER
REACTOR





VI0707-17590 ON 540

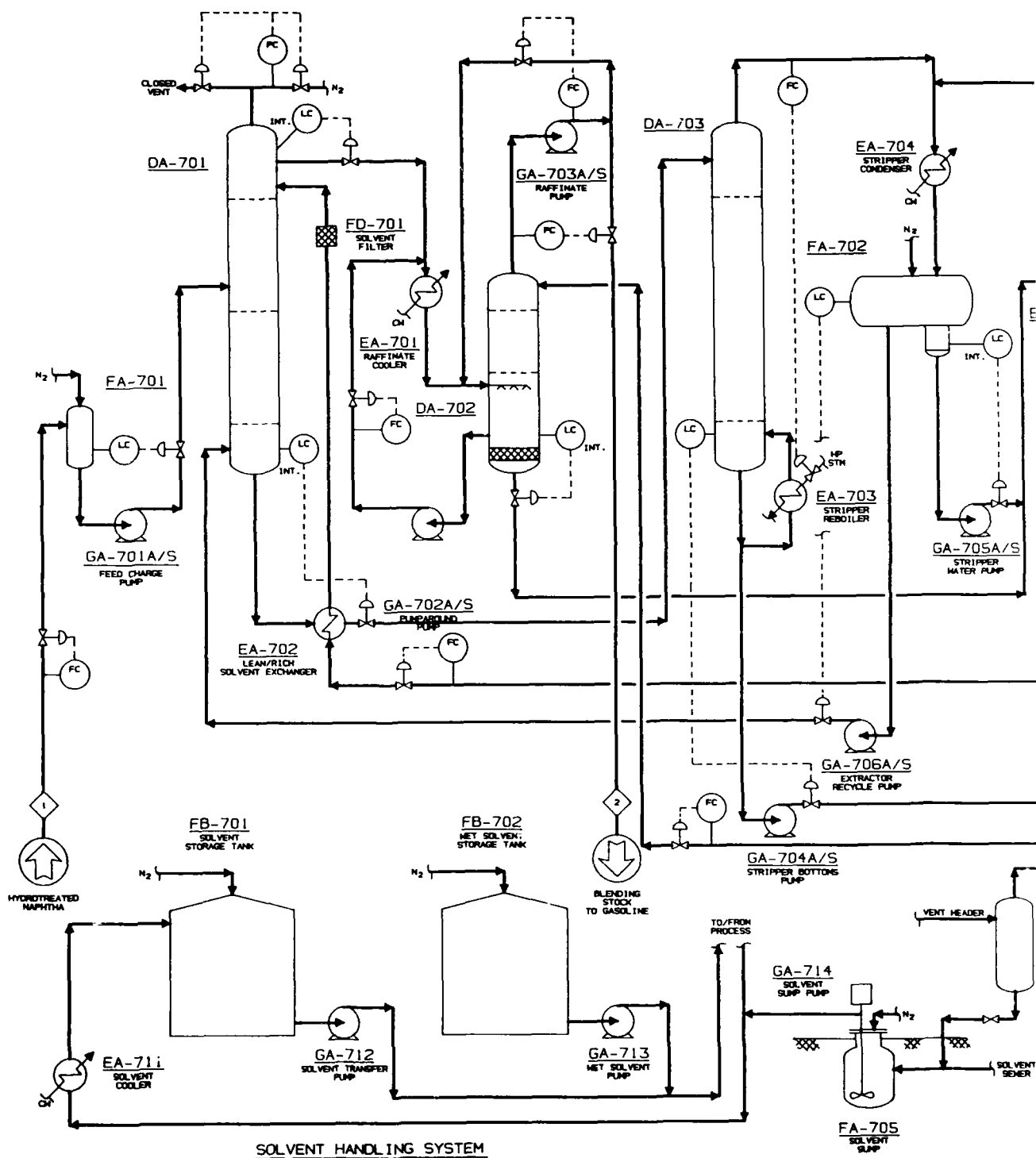
FA-701
FEED SURGE
DRUM

DA-701
EXTRACTOR
COLUMN

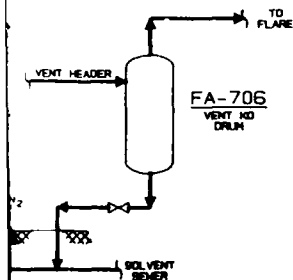
DA-702
RAFFINATE
WASH
COLUMN

DA-703
STRIPPER

FA-702
STRIPPER
REFLUX DRUM



FA-704
EJECTOR CONDENSATE
DRUM



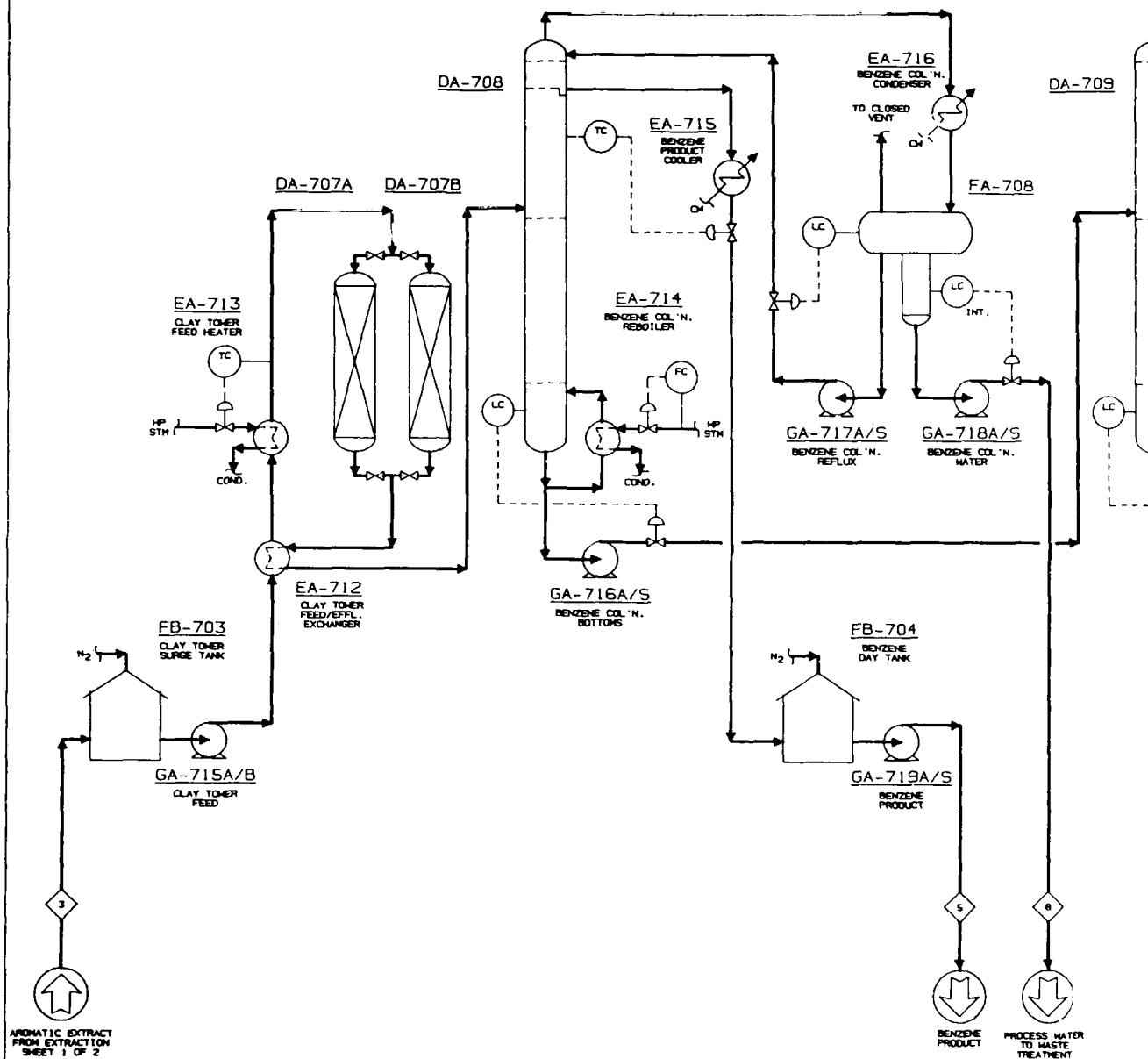
SCALE	ENG. NO.	D5571-70701A

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CLAY TOWERS BENZENE COLUMN

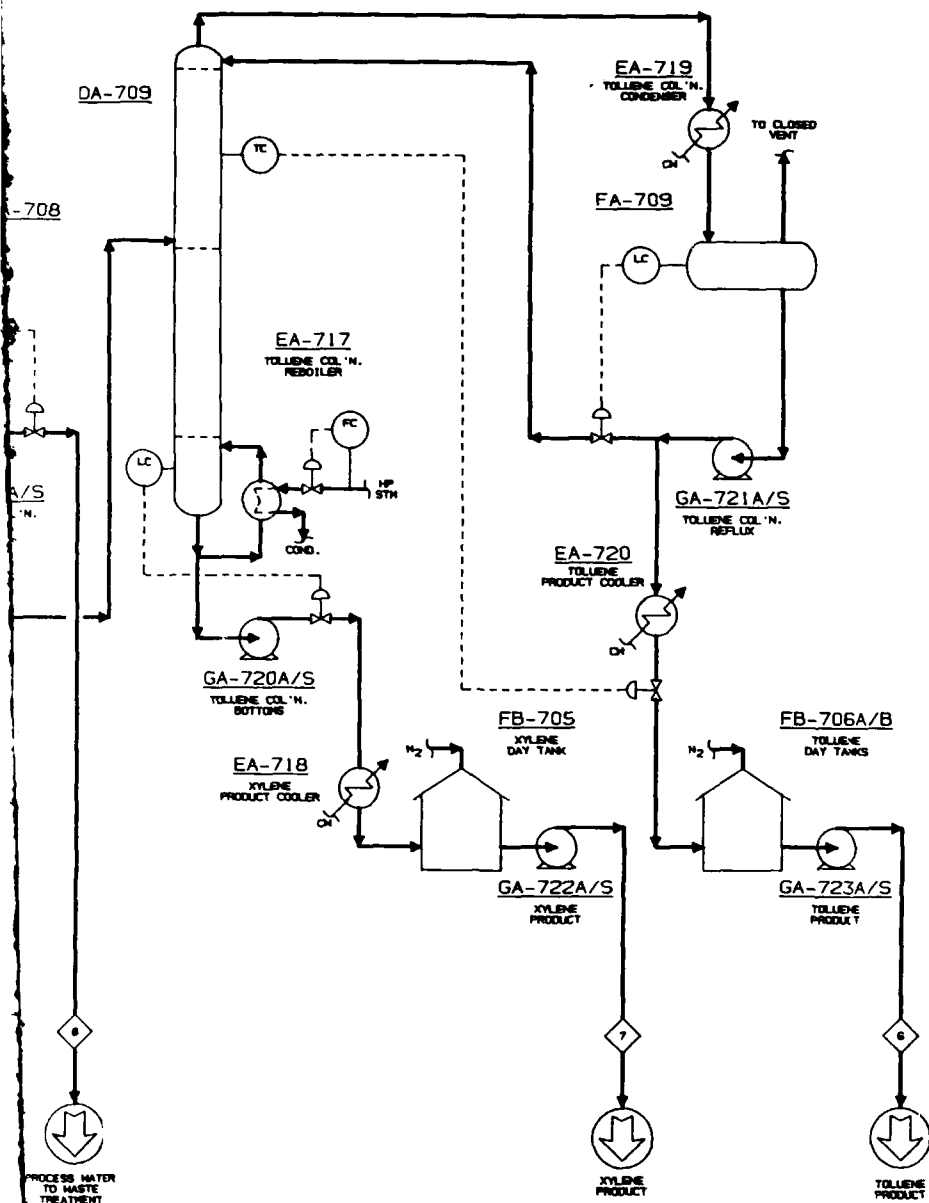
BENZENE COLUMN
REFLUX DRUM

TOLUEN



DA-709
TOLUENE COLUMN

FA-709
TOLUENE COLUMN
REFLUX DRUM



SHELL SULFOLANE PROCESS

AMOCO/DOE - JET FUELS FROM COAL
DERIVED LIQUIDS

REV.	DATE	DESCRIPTION	BY	CHKD	APP	DATE	REV.	DATE	DESCRIPTION	BY	CHKD	APP	DATE

COMBUSTION ENGINEERING

LUMPLINE CHEM INC.
Bloomfield, NJ

PROCESS FLOW DIAGRAM
AROMATICS RECOVERY UNIT
AROMATICS FRACTIONATION SECTION
AREA 700 CASE 7 SHEET 2 OF 2

SCALE DWS. NO. DS571-70701B

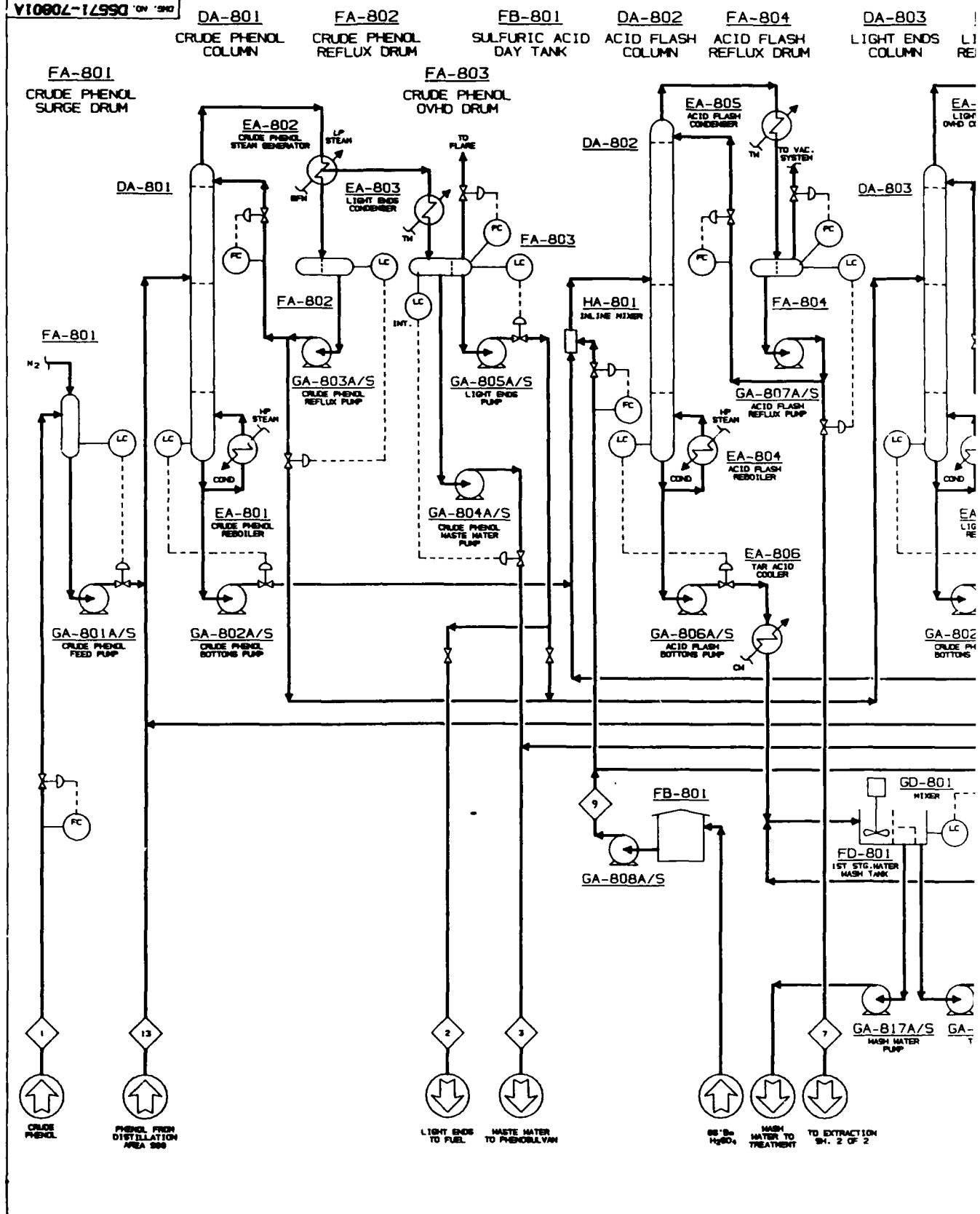


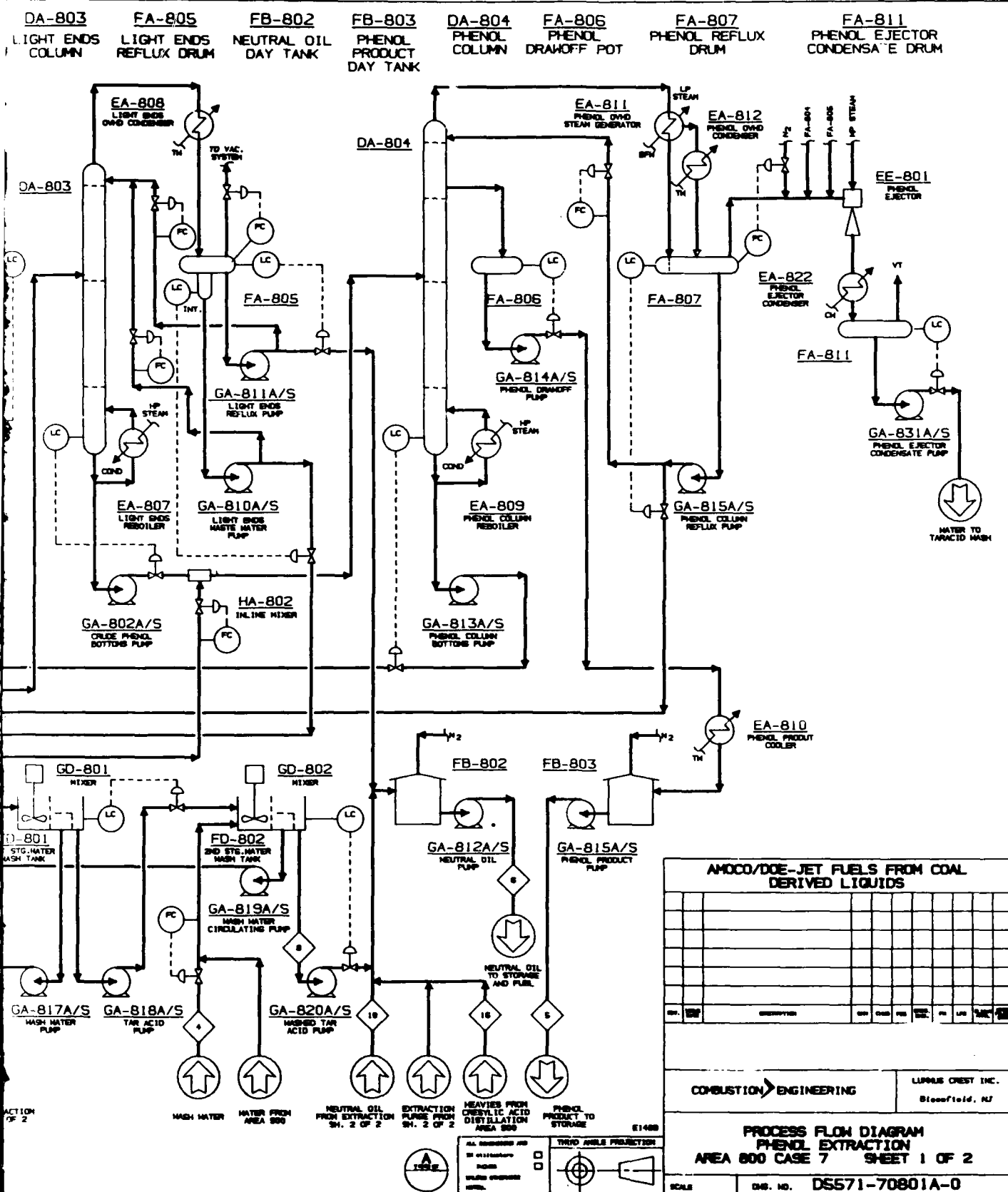
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IN MILLIMETERS
UNLESS OTHERWISE
SPECIFIED

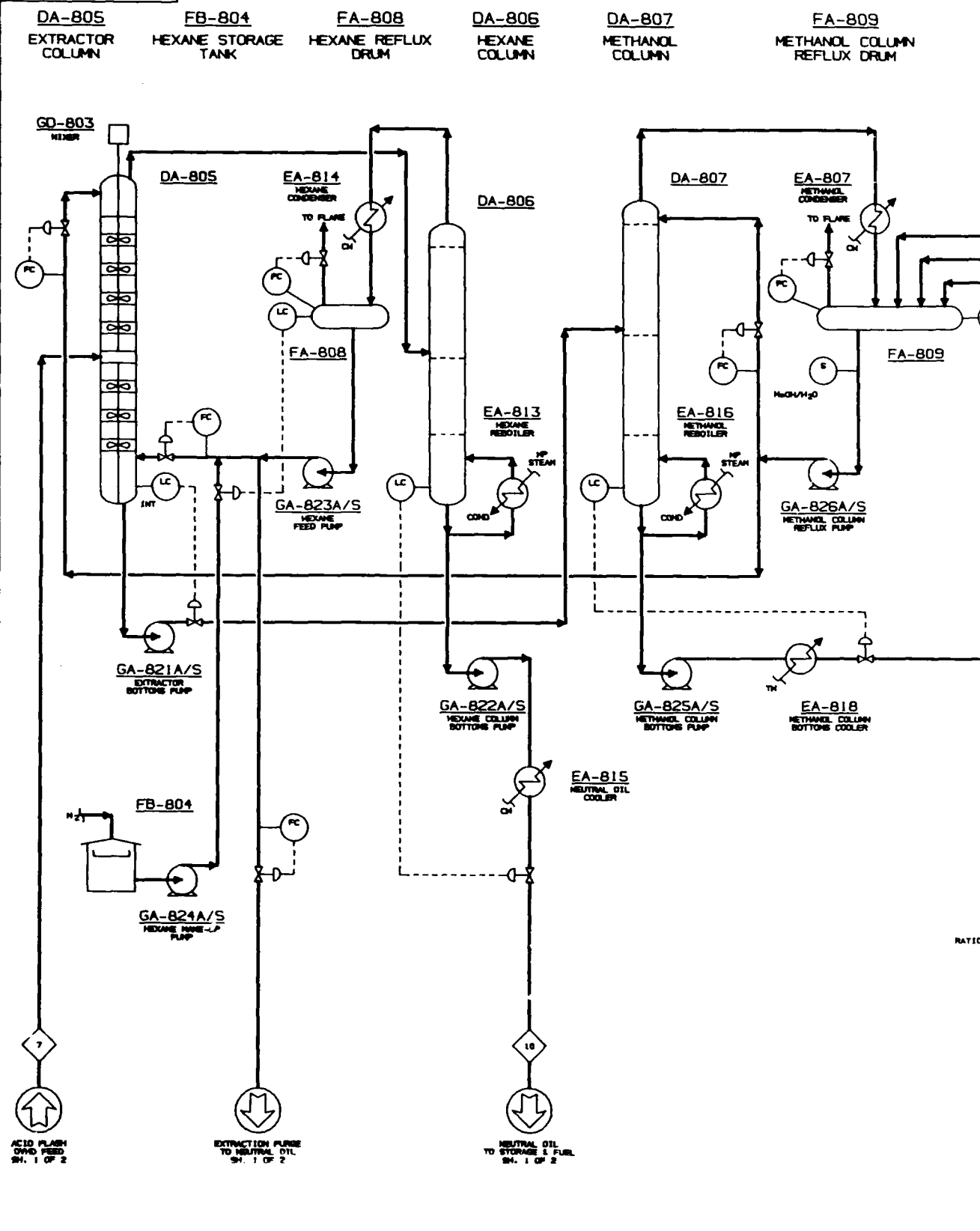
THIRD ANGLE PROJECTION
E1405

B

Y10802-17590 ON '802



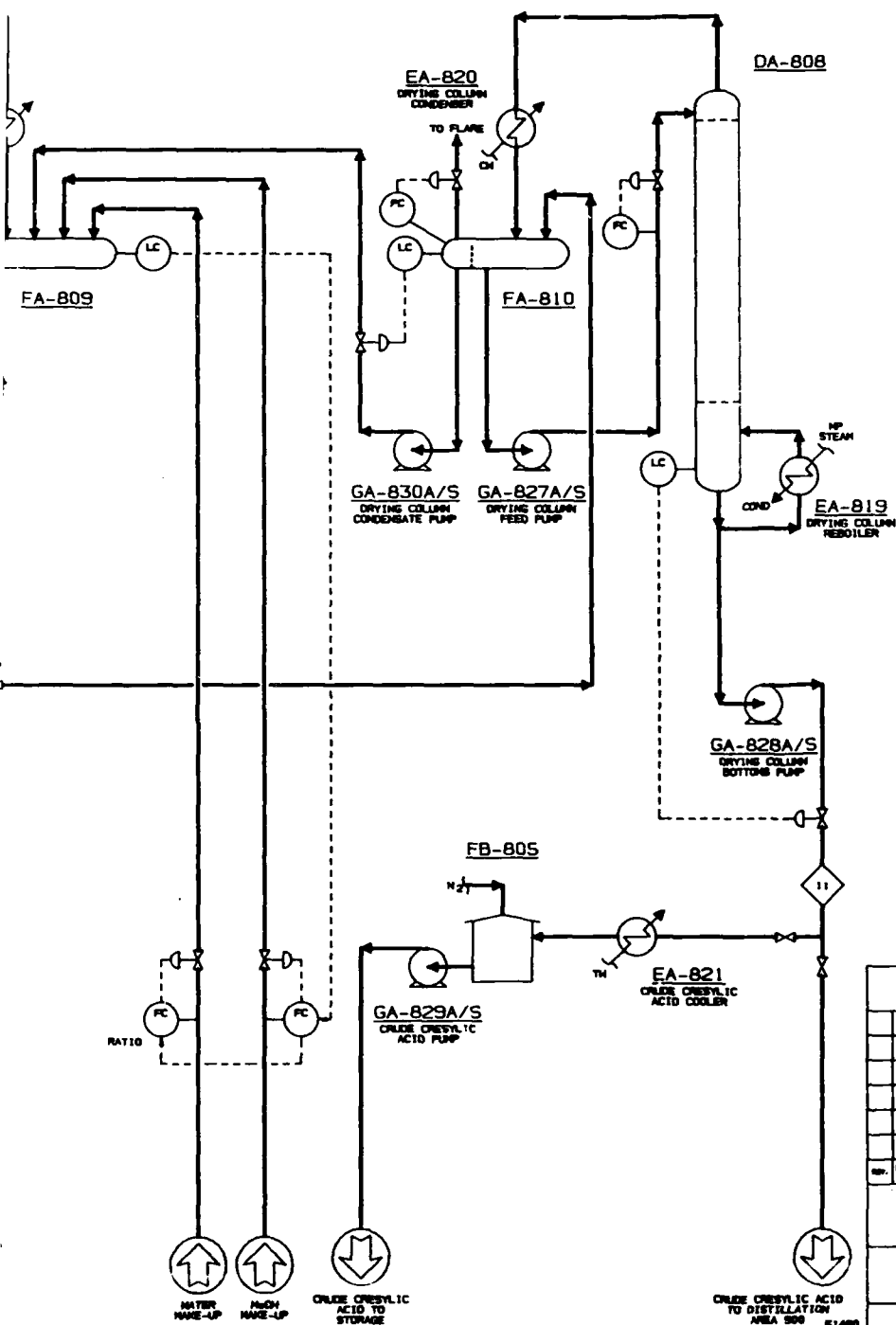




FB-805
CRUDE CRESYLIC
ACID DAY TANK

FA-810
DRYING COLUMN
REFLUX DRUM

DA-808
DRYING COLUMN



AMOCO/DOE-JET FUELS FROM COAL
DERIVED LIQUIDS

[illegible]

COMBUSTION ENGINEERING

LUNDS CREST INC.
Blacksfield, NJ

PROCESS FLOW DIAGRAM
PHENOL EXTRACTION
AREA 800 CASE 7 SHEET 2 OF 2

SCALE

DWG. NO. D5571-70801B-0

ALL INFORMATION CONTAINED
HEREIN IS UNCLASSIFIED
DATE 08-22-2001 BY 60322
UCBAW

THIRD ANGLE PROJECTION



10607-1/553

DA-901
O-CRESOL
COLUMN

DA-902
O-CRESOL
STRIPPER

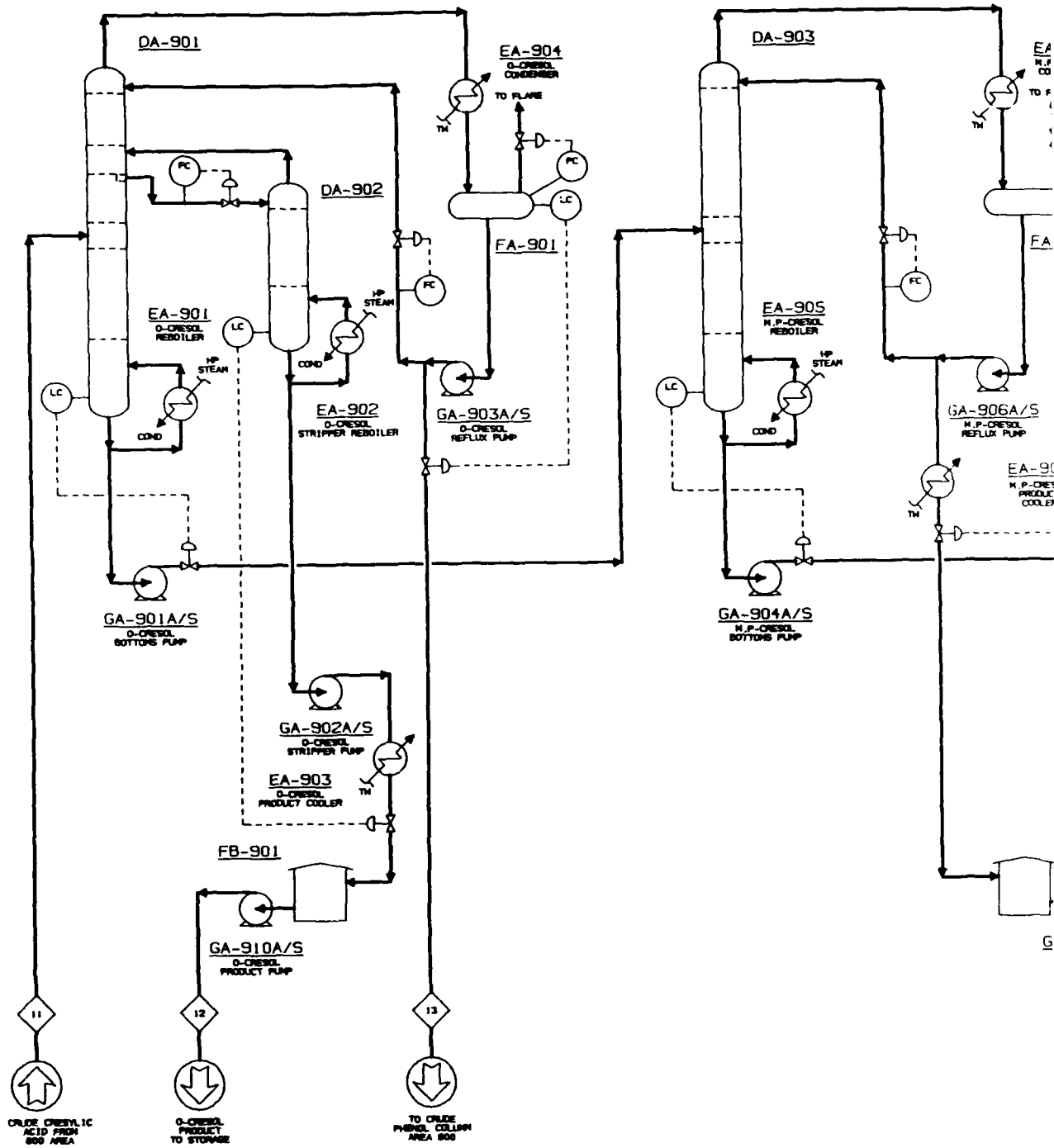
FB-901
O-CRESOL
DAY TANK

FA-901
O-CRESOL
REFLUX DRUM

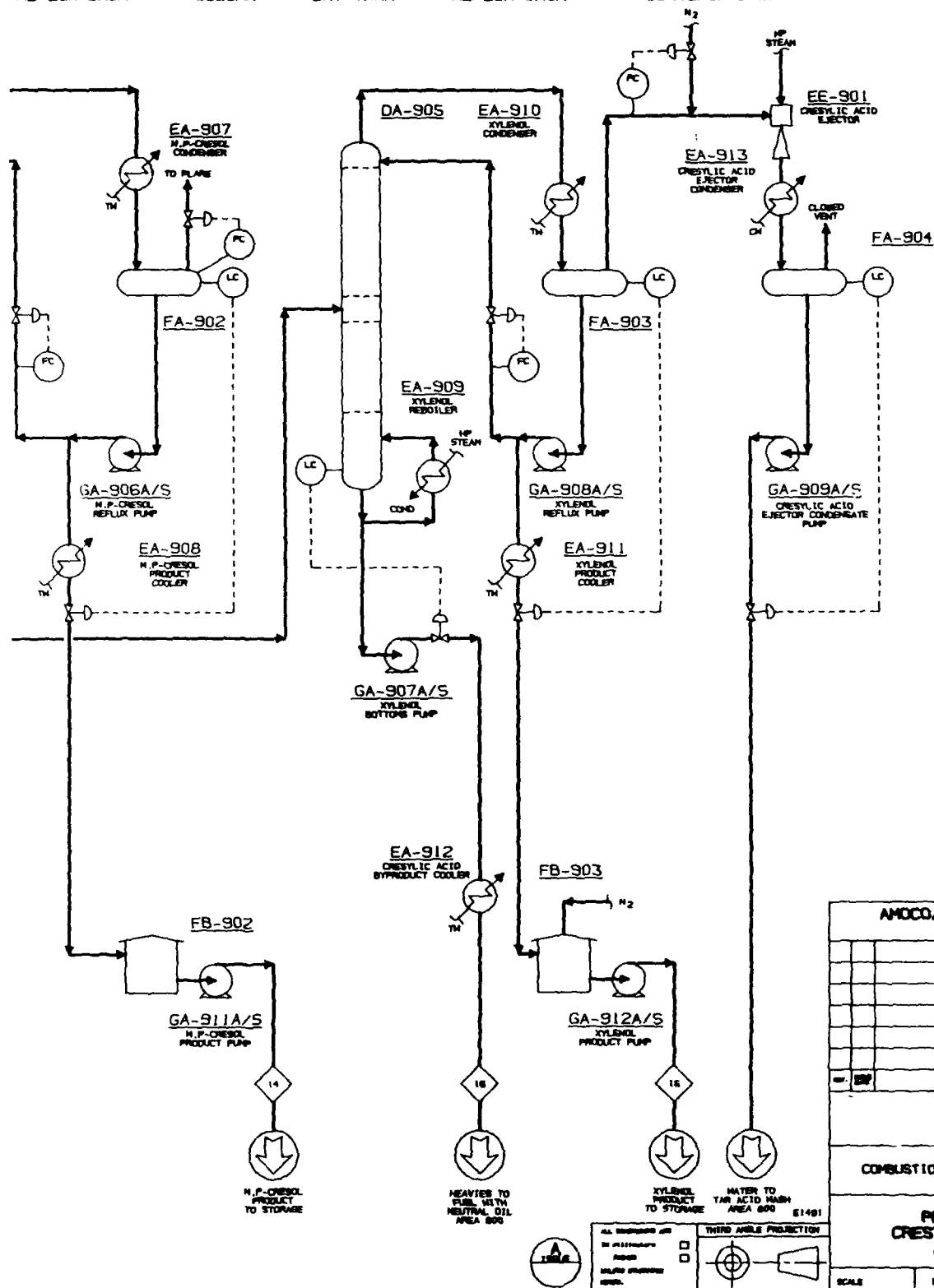
DA-903
M,P-CRESOL
COLUMN

FB-902
M,P-CRESOL
DAY TANK

FA-902
M,P-CRESOL
REFLUX DRUM



FA-904
CRESYLIC ACID EJECTOR
CONDENSATE DRUM

[illegible]

3.0 CAPITAL COSTS

3.1 Phenol Stream

3.1.1 Equipment List

Case 7 Maximum Profit

Area 400 - Storage Area

FB-410	Neutral Oil Storage
FB-411	Phenol Product Storage
FB-412	Crude Cresylic Acid Storage
FB-413	O-Cresol Storage
FB-414	M, P Cresol Storage
FB-415	Xylenol Storage
GA-402A/S	Crude Phenol Feed Pump
GA-412A/S	Neutral Oil Transfer Pump
GA-413A/S	Crude Cresylic Acid Transfer Pump
GA-414A/S	O-Cresol Transfer Pump
GA-415A/S	M, P Cresol Transfer Pump
GA-416A/S	Xylenol Transfer Pump

Area 800 - Phenol Extraction

DA-801	Crude Phenol Column
DA-802	Acid Flash Column
DA-803	Light Ends Column
DA-804	Phenol Column
DA-805	Extractor Column
DA-806	Hexane Column
DA-807	Methanol Column
DA-808	Drying Column
EA-801	Crude Phenol Reboiler
EA-802	Crude Phenol Steam Generator
EA-803	LT. Ends Condenser
EA-804	Acid Flash Reboiler
EA-805	Acid Flash Condenser
EA-806	Tar Acid Cooler
EA-807	LT. Ends Reboiler
EA-808	LT. Ends OVHD Condenser
EA-809	Phenol Reboiler
EA-810	Phenol Product Cooler
EA-811	Phenol OVHD Stm. Generator
EA-812	Phenol OVHD Condenser
EA-813	Hexane Reboiler
EA-814	Hexane Condenser
EA-815	Neutral Oil Cooler
EA-816	Methanol Reboiler

3.1 Phenol Stream

3.1.1 Equipment List - cont'd

<u>Area - 800</u>	<u>Phenol Extractor</u>
EA-817	Methanol Condenser
EA-818	Methanol Column Bottoms Cooler
EA-819	Drying Column Reboiler
EA-820	Drying Column Condenser
EA-821	Crude Cresylic Acid Cooler
EA-822	Phenol Ejector Condenser
EE-801	Phenol Ejector
FA-801	Crude Phenol Surge Drum
FA-802	Crude Phenol Reflux Drum
FA-803	Crude Phenol OVHD Drum
FA-804	Acid Flash Reflux Drum
FA-805	Light Ends Reflux Drum
FA-806	Phenol Drawoff Pot
FA-807	Phenol Reflux Drum
FA-808	Hexane Reflux Drum
FA-809	Methanol Reflux Drum
FA-810	Drying Column Reflux Drum
FA-811	Phenol Ejector Condensate Drum
FB-801	Sulfuric Acid Day Tank
FB-802	Neutral Oil Day Tank
FB-803	Phenol Product Day Tank
FB-804	Hexane Storage Tank
FB-805	Crude Cresylic Acid Day Tank
FD-801	1st Stg. Water Wash Tank
FD-802	2nd Stg. Water Wash Tank
GA-801A/S	Crude Phenol Feed Pump
GA-802A/S	Crude Phenol Bottoms Pump
GA-803A/S	Crude Phenol Reflux Pump
GA-804A/S	Crude Phenol Waste Water Pump
GA-805A/S	Light Ends Pump
GA-806A/S	Acid Flash Bottoms Pump
GA-807A/S	Acid Flash Reflux Pump
GA-808A/S	Sulfuric Acid Feed Pump
GA-809A/S	Light Ends Bottoms Pump
GA-810A/S	Light Ends Waste Water Pump
GA-811A/S	Light Ends Reflux Pump
GA-812A/S	Neutral Oil Pump
GA-813A/S	Phenol Column Bottoms Pump
GA-814A/S	Phenol Drawoff Pump
GA-815A/S	Phenol Product Pump
GA-816A/S	Phenol Column Reflux Pump

3.1 Phenol Stream

3.1.1 Equipment List - cont'd

Area 800 - Phenol Extractor

GA-817A/S	Wash Water Pump
GA-818A/S	Acid Tar Pump
GA-819A/S	Wash Water Circulating Pump
GA-820A/S	Washed Tar Acid Pump
GA-821A/S	Extractor Bottoms Pump
GA-822A/S	Hexane Column Bottoms Pump
GA-823A/S	Hexane Feed Pump
GA-824A/S	Hexane Make-up Pump
GA-825A/S	Methanol Column Bottoms Pump
GA-826A/S	Methanol Column Reflux Pump
GA-827A/S	Drying Column Feed Pump
GA-828A/S	Drying Column Bottoms Pump
GA-829A/S	Crude Cresylic Acid Pump
GA-830A/S	Drying Column Condensate Pump
GA-831A/S	Phenol Ejector Condensate Pump

GD-801	1st Stg. Water Wash Mixer
GD-902	2nd Stg. Water Wash Mixer
GD-803	Extractor Mixer

HA-801	Inline Mixer
HA-802	Inline Mixer

Area 900 - Cresylic Acid Distillation

DA-901	O-Cresol Column
DA-902	O-Cresol Stripper
DA-903	M,P-Cresol Column
DA-905	Xylenol Column

EA-901	O-Cresol Reboiler
EA-902	O-Cresol Stripper Reboiler
EA-903	O-Cresol Product Cooler
EA-904	O-Cresol Condenser
EA-905	M,P-Cresol Reboiler
EA-907	M,P-Cresol Condenser
EA-908	M,P-Cresol Product Cooler
EA-909	Xylenol Reboiler
EA-910	Xylenol Condenser
EA-911	Xylenol Product Cooler
EA-912	Cresylic Acid By-product Cooler
EA-913	Cresylic Acid Ejector Condenser

EE-901	Cresylic Acid Ejector

3.1 Phenol Stream

3.1.1 Equipment List - cont'd

Area 900 - Cresylic Area Distillation

FA-901	O-Cresol Reflux Drum
FA-902	M,P-Cresol Reflux Drum
FA-903	Xylenol Reflux Drum
FA-904	Cresylic Acid Ejector Condensate Drum
FB-901	O-Cresol Day Tank
FB-902	M,P-Cresol Day Tank
FB-903	Xylenol Day Tank
GA-901A/S	O-Cresol Bottoms Pump
GA-902A/S	O-Cresol Stripper Pump
GA-903A/S	O-Cresol Reflux Pump
GA-904A/S	M,P-Cresol Bottoms Pump
GA-906A/S	M,P-Cresol Reflux Pump
GA-907A/S	Xylenol Bottoms Pump
GA-908A/S	Xylenol Reflux Pump
GA-909A/S	Cresylic Acid Ejector Cond. Pump
GA-910A/S	O-Cresol Product Pump
GA-911A/S	M,P-Cresol Product Pump
GA-912A/S	Xylenol Product Pump

3.1 Phenol Stream - cont'd

3.1.2 Cost Estimate

3.1.2.1 Basis of Estimate

The estimate is an equipment factored type estimate using the equipment sizes & specifications developed for this project. The equipment unit pricing is based on return data for equipment purchased for various projects. The unit pricing is some what conservative compared to world wide markets of 2-3 years ago, however, the exchange rate decline during this period will lead to higher purchase prices.

The commodity materials & subcontracts are ratioed from the equipment costs using factors considering the process, the size of the units, and the location of the plant.

The labor and indirects also are factored considering process, sizing, and location.

Engineering costs are based on the equipment count times the historical number of manhours per equipment item, and the current average engineering selling rate.

In light of the preliminary process basis developed for this section of the plant, a 30% contingency has been applied to the base costs.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.1.2.2 Estimate Summary

(Thousands of \$)

Case 7

Area 800 Phenol Extraction	12276
Area 900 Cresylic Acid Distil.	4832
Area 400 OSBL (Phenol Stream Part)	<u>3016</u>
Total	\$20,124

3.1 Phenol Stream - cont'd

3.1.2.3 Estimate Breakdown

Area 800

<u>Equipment</u>		<u>\$ Value</u>	<u>% Comm.</u>	<u>\$ Comm.</u>
<u>Items</u>	<u>Type</u>			
8	Towers	865	50	432
-	Internals	90	-	-
22	Exchangers	375	100	375
11	Vessels	151	120	181
5	Tanks	74	120	89
2	Filters	30	100	30
62	Pumps	353	120	424
Total	110	\$2038		\$1531

Equipment 2038
Commodities 1531
Labor 1122 10% Equip, 60% Comm.
Indirects 1122
Office 3630 110 pcs. x 600 x \$55
Subtotal \$9,443

Contingency 2833 30%
\$12,276

Area 900

<u>Equipment</u>		<u>\$ Value</u>	<u>% Comm.</u>	<u>\$ Comm.</u>
<u>Items</u>	<u>Type</u>			
4	Towers	219	120	263
-	Internals	68	-	-
12	Exchangers	138	110	152
4	Vessels	42	120	50
3	Tanks	20	120	24
24	Pumps	132	120	158
Total	47	\$ 619		\$ 647

Equipment 619
Commodities 647
Labor 450 10% Equip, 60% Comm.
Indirects 450
Office 1551 47 x 600 x \$55
Subtotal \$3,717

Contingency 1,115 30%
\$4,832

3.1 Phenol Stream - cont'd

3.1.2.3 Estimate Breakdown - cont'd

Area 400

Tankage	MTL	S/C	\$199
Pumps	MTL		54
Equipment related commodities			200
Yard Piping 29000 LF	MTL		\$330
Labor			870
Pipe Insul.	S/C		178
Tracing	S/C		132
Excavation	S/C		150
Rack	S/C		<u>300</u>
Subtotal			\$2,413
Contingency	25%		<u>603</u>
Total			\$3,016

AMOCO/DOE
GREAT PLAINS GASIFICATION PLANT
JET FUEL FROM COAL DERIVED LIQUIDS

3.0 CAPITAL COSTS

3.2 Naphtha Stream

3.2.1 Equipment List

Case 7 - Maximum Profit

Area 300 - PSA Hydrogen Unit

<u>Tag. No.</u>	<u>Description</u>
PA-301	PSA Hydrogen Unit Package

Area 400 - Storage

FB-403	Fuel Oil Storage
FB-404	Blending Stock
FB-405	Benzene Storage
FB-406	Toluene Storage
FB-407	Xylene Storage
FB-408	Butane Storage
FB-409	Gasoline Storage

GA-403A/S	Fuel Oil Transfer Pump
GA-405A/S	Crude Naphtha Pump
GA-406A/S	Blending Stock Pump
GA-407A/S	Benzene Transfer Pump
GA-408A/S	Toluene Transfer Pump
GA-409A/S	Xylene Transfer Pump
GA-410A/S	Butane Transfer Pump
GA-411A/S	Gasoline Transfer Pump

PA-401 Gasoline Blending Package

Area 600 - Naphtha Distillation & HDT

DA-601	Naphtha Distillation Column
DA-602	HDT Naphtha Stabilizer Column
DC-601	Naphtha Hydrotreater Reactor

3.2 Naphtha Stream

3.2.1 Equipment List - cont'd

Area 600 - Naphtha Distillation & HDT

EA-601	Naphtha Distillation Column Reboiler
EA-602	Naphtha Distillation Column Condenser
EA-603	HDT Reactor Feed/Effl. Exchanger
EA-604	HDT Reactor Recycle Gas Heater
EA-605	Stabilizer Feed Exchanger
EA-606	Reactor Effl. Cooler
EA-607	Stabilizer Reboiler
EA-608	Naphtha Stabilizer Condenser
EA-609	HDT Naphtha Cooler

FA-601	Crude Naphtha Feed Surge Drum
FA-602	Distil. Col'n Reflux Drum
FA-603	HDT Feed Surge Drum
FA-604	Makeup Gas KO Drum
FA-606	LT/HP Separator
FA-607	Recycle Gas KO Drum
FA-608	Naphtha Stabil. Reflux Drum

GA-601A/S	Crude Naphtha Feed Pump
GA-602A/S	Distil. Col'n Bottoms Pump
GA-603A/S	Distil. Col'n Reflux Pump
GA-604A/S	HDT Feed Pump
GA-605A/S	Process Water Pump
GA-606A/S	HDT Naphtha Pump
GA-607A/S	Naphtha Stabil. Reflux Pump
GA-608A/S	Sour Water Pump

GB-601A/B	Makeup Gas Compressor
GB-602A/B	Recycle Gas Compressor

3.2 Naphtha Stream

3.2.1 Equipment List - cont'd

Area 700 - Aromatics Recovery

DA-701	Extractor Column
DA-702	Raffinate Water Wash Column
DA-703	Stripper
DA-704	Recovery Column
DA-705	Water Stripper
DA-706	Solvent Regenerator
DA-707A/B	Clay Tower
DA-708	Benzene Column
DA-709	Toluene Column
EA-701	Raffinate Cooler
EA-702	Lean/Rich Solvent Exchanger
EA-703	Stripper Reboiler
EA-704	Stripper Condenser
EA-705	Recovery Column Reboiler
EA-706	Recovery Column Intermediate Reboiler
EA-707	Recovery Column Condenser
EA-708	Recovery Column Ejector Condenser
EA-709	Water Stripper Reboiler
EA-710	Solvent Regenerator Reboiler
EA-711	Solvent Cooler
EA-712	Clay Tower Feed/Effl. Exchanger
EA-713	Clay Tower Feed Heater
EA-714	Benzene Column Reboiler
EA-715	Benzene Product Cooler
EA-716	Benzene Column Condenser
EA-717	Toluene Column Reboiler
EA-718	Xylene Product Cooler
EA-719	Toluene Column Condenser
EA-720	Toluene Product Cooler
EE-701	Recovery Column Ejector
FA-701	Feed Surge Drum
FA-702	Stripper Reflux Drum
FA-703	Recovery Column Reflux Drum
FA-704	Ejector Condensate Drum
FA-705	Solvent Sump
FA-706	Vent KO Drum
FA-708	Benzene Column Reflux Drum
FA-709	Toluene Column Reflux Drum

3.2 Naphtha Stream

3.2.1 Equipment List - cont'd

FB-701	Solvent Storage Tank
FB-702	Wet Solvent Storage Tank
FB-703	Clay Tower Surge Tank
FB-704	Benzene Day Tank
FB-705	Xylene Day Tank
FB-706A/B	Toluene Day Tanks
FD-701	Solvent Filter
GA-701A/S	Feed Charge Pump
GA-702A/S	Pumparound Pump
GA-703A/S	Raffinate Pump
GA-704A/S	Stripper Bottoms Pump
GA-705A/S	Stripper Water Pump
GA-706A/S	Extractor Recycle Pump
GA-707A/S	Lean Solvent Pump
GA-708A/S	Wash Water Pump
GA-709A/S	Recovery Column OV'HD Pump
GA-710A/S	Water Stripper Bottoms Pump
GA-711A/S	Ejector Condensate Pump
GA-712	Solvent Transfer Pump
GA-713	Wet Solvent Pump
GA-714A/B	Solvent Sump Pump (Warehouse Spare)
GA-715A/S	Clay Tower Feed Pump
GA-716A/S	Benzene Column Bottoms Pump
GA-717A/S	Benzene Column Reflux Pump
GA-718A/S	Benzene Column Water Pump
GA-719A/S	Benzene Product Pump
GA-720A/S	Toluene Column Bottoms Pump
GA-721A/S	Toluene Column Reflux Pump
GA-722A/S	Xylene Product Pump
GA-723A/S	Toluene Product Pump
PA-701	Clay Handling Equipment

3.2.2 Cost Estimate

3.2.2.1 Basis of Estimate

The estimate is an equipment factored type estimate using the equipment sizes & specifications developed for this project. The equipment unit pricing is based on return data for various projects. The unit pricing is somewhat conservative compared to world wide markets of 2-3 years ago, however, the exchange rate decline during this period will lead to higher purchase prices.

3.2.2 Cost Estimate

3.2.2.1 Basis of Estimate - cont'd

The commodity materials & subcontracts are ratioed from the equipment costs using factors considering the high pressure processing, the size of the units, and the location of the plant.

The labor and indirects also are factored considering process, sizing, and location.

Engineering costs are based on the equipment count times the historical number of manhours per equipment item, and the current average engineering selling rate.

In light of the preliminary data developed for this effort, a 20% contingency has been applied to the base costs.

Excluded from this estimate are:

Spare Parts
Start-Up
Insurances & Taxes
Permits
Royalties on Processing Technology Knowhow

3.2.2.2 Estimate Summary

(Thousands of \$)

	<u>Case 7</u>
Area 600 Naphtha Distill. & HDT	4615
Area 700 Aromatics Recovery	9373
Area 300 PSA	518
Area 400 OSBL (Naphtha Stream Part)	<u>3058</u>
Subtotal	\$17,564
Area 700 - ARU Solvent Inventory	<u>100</u>
Total	\$17,664

3.2 Naphtha Stream - cont'd

3.2.2.3 Estimate Breakdown (Area 600) All values in Thousands

<u>Equipment</u>		<u>\$ Value</u>	<u>% Comm.</u>	<u>\$ Comm.</u>
<u>Items</u>	<u>Type</u>			
2	Towers	48	140	67
-	Internals	8	-	-
1	Exchangers	125	85	106
9	Vessels	123	100	123
7	Tanks	89	100	89
16	Pumps	68	100	68
4	Compressors	230	60	138
Total	39	\$ 691		\$ 591

Equipment 691
Commodities 591
Labor 424 10% Equip., 60% Comm.
Indirects 424
Office 1716 39 pcs. x 800 x \$55
Subtotal \$3,846

Contingency 769 20%
Total \$4,615

Area 700

<u>Equipment</u>		<u>\$ Value</u>	<u>% Comm.</u>	<u>\$ Comm.</u>
<u>Items</u>	<u>Type</u>			
10	Towers	350	140	490
8	Internals	66	-	-
20	Exchangers	113	100	113
9	Vessels	65	120	78
7	Tanks	117	100	117
44	Pumps	180	120	216
3	Special	20	100	20
Total	101	\$ 911		\$1034

Equipment 911
Commodities 1034
Labor 711 10% Equip., 60% Comm.
Indirects 711 100%
Office 4444 101 pcs. x 800 x \$55
Subtotal \$7,811

Contingency 1,562 20%
Total \$9,373

3.2 Naphtha Stream - cont'd

3.2.2.3 Estimate Breakdown (Area 600) All Values in Thousands
- cont'd

Area 300

PSA Hydrogen Package Unit (one skid)

205 MMSCFD budget quote -	\$300
Installation 50%	<u>150</u>
Subtotal	\$450
Contingency 15%	<u>68</u>
Total	\$518

Area 400

Equipment & Value

Tankage	MTL	S/C	366
Pumps	MTL		62
Yard Piping 27,000 LF	MTL		300
Labor	S/C		813
Pipe Insulation	S/C		108
Tracing	S/C		84
Excavation	S/C		135
Rack	S/C		300
Equipment Related Commodities 65% Equip.			278
Subtotal			<u>\$2,446</u>
Contingency 25%			<u>612</u>
			<u>\$3,058</u>

4.0 OPERATING COSTS

4.1 Phenol Stream

4.1.1 Operating Labor

It is estimated that it will require 6 men/shift to operate the plant broken down as follows:

Foreman	1
Control Room	1
Phenol Extr. Operator	1
Crsylic Acid Distill. Operator	1
	<u>4</u>
	Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 3 people.

The total additional people (assuming 4 & 2 operation for the process units) are as follows:

Shift Personnel	4 positions x 4 people/ position -	16
Supervisor & Admin.		5
QC Technician		1
Maintenance		3
Other (Stores or Janitorial)		1
Total		<u>26</u>

4.1.2 Utilities

The following utilities have been estimated from the preliminary process designs:

<u>Utility</u>	<u>Consumption</u>	<u>Cost</u>	<u>\$/SD</u>
#6 Fuel Oil	626 BPSD	\$16/Bbl (a)	10,016
Cooling Water	2300 GPM	\$0.155/MGal (b)	513
Power	290 kW	\$0.04/kWH (b)	278
Process Water	2 GPM	\$0.45/MGal (b)	2

(a) Cut of 1% sulfur \$6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

(b) ANG utility cost information dated 5/87.

4.1.3 Chemicals

The chemical cost is as follows:

<u>Chemical</u>	<u>Use</u>	<u>Cost</u>	<u>\$/SD</u>
H ₂ SO ₄	7100 #/D	\$0.04/#	285

4.1.4 Maintenance Supplies

Maintenance supplies costs are not known but will be assumed to be 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units. On this basis the maintenance supplies would be

$$0.00005 \times 20,124,000 = \$1006/\text{SD}$$

4.2 Naphtha Stream

4.2.1 Operating Labor

It is estimated that it will require 7 men/shift to operate this section plant broken down as follows:

Foreman	1
Control Room	1
HDT Operator	2
ARU Operator	2
PSA & relief man	1
	<u>7</u> Shift Positions

Maintenance will probably be provided from the existing SNG plant maintenance shop. It is likely that the existing maintenance department will be expanded by about 5 people.

The total additional people (assuming 6 & 2 operation for the process units) are as follows:

Shift Personnel	7 positions x 4 people/position -	28
Supervisor & Admin.		5
QC Technician		1
Maintenance		5
Other (Stores or Janitorial)		1
Total		<u>40</u>

4.2.2 Utilities

The following utilities have been estimated from the simulations:

<u>Utility</u>	<u>Consumption</u>	<u>Cost</u>	<u>\$/SD</u>
#6 Fuel Oil	567 BPSD	\$16/Bbl (a)	9072
SNG equivalent	0.17 MM SCFD	\$3.80/MM BTU (b)	633
Cooling Water	1700 GPM	\$0.155/MGal (c)	380
Power	235 kW	\$0.04/kWH (c)	226
Process Water	1 GPM	\$0.45/MGal (c)	1

(a) Cut of 1% sulfur \$6 oil in Minnesota on 11/24/87 as per Platts Oilgram.

(b) Memo from D. Daley of BRSC to L. Lorenzo of DOE dtd. 10/20/87, Ref. Dpd-87-863.

(c) ANG utility cost information dated 5/87.

4.2.3 Catalyst & Chemicals

The catalyst and chemicals cost is as follows:

<u>Catalyst & Chem</u>	<u>Use</u>	<u>Cost</u>	<u>\$/SD</u>
HDT Cat.	0.021 #/Bbl	\$3.00/#	33
ARU Solvent	24 #/D	\$2.10/#	50
			\$88

4.2.4 Maintenance Supplies

Maintenance supplies for refining operations typically cost between 1.5-2.0% of the installed cost per year. For a daily cost we would estimate the cost of maintenance supplies to be 0.005% of the total installed cost of the process units excluding solvent inventory. On this basis the maintenance supplies would be

$$0.00005 \times 17,564,000 = \$878/\text{SD}$$

5.0 PLOT PLAN AND UNIT TIE-INS

5.1 Plot Plan

The process units required for the maximum profit case are proposed to be located to the east of the Phenosolvan and Waste Water Units of the existing gasification plant as indicated on the markup of the overall Process Area Plot Plan, LCI Dwg E7102-00010A. This area approx. 375' x 350' will be surrounded by an access road and will be divided by a central east-west road. Areas 300, 600 & 700 will be located to the south and Areas 800 & 900 to the north of the road.

A diked storage tank area approx. 360' x 260' will be required for product and fuel oil storage associated with the naphtha stream and an area approx. 60' x 300' is required for the phenol stream. These areas are proposed to be located to the south of the existing tankage area adjacent to the railcar loading spurs.

5.2 Unit Tie-Ins

Approximately 1200 ft of new interconnecting pipe rack will be required to connect the new process area with the main yard rack of the gasification plant, the product storage area and flare.

New storm, oily water and sanitary sewer lines will be run from the new process units south to their respective collection systems.

A summary of the lines is shown in table 5.1.

TABLE 5.1
INTERCONNECTING PIPING

I - PHENOL STREAM

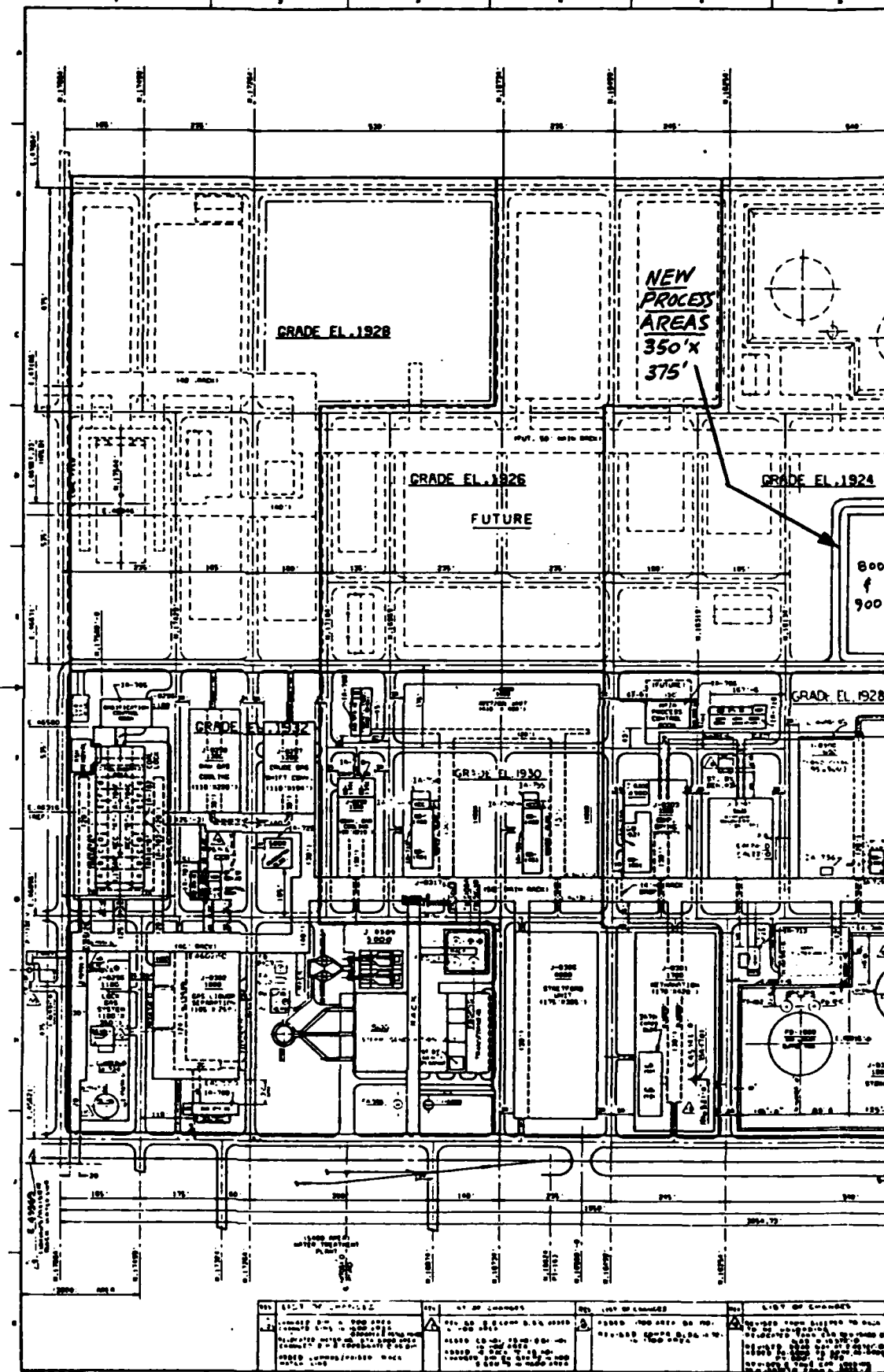
<u>EST. SIZE</u>	<u>SERVICE</u>	<u>TO/FROM</u>
2"	Crude Phenol (Elec. Tr.)	Storage
1 1/2"	Light Ends	Storage/Ph. Ext.
1 1/2"	Neutral Oil	Storage
1 1/2"	Phenol Product (Elec. Tr.)	Storage
1 1/2"	Crude Cresylic Acid (Elec. Tr.)	Storage
1 1/2"	O-Cresol (Elec. Tr.)	Storage
1 1/2"	M,P-Cresol (elec. Tr.)	Storage
1 1/2"	Xylenol Product (Elec. Tr.)	Storage
1 1/2"	Methanol Make-up	Ph. Ext./MeoH Unit
1 1/2"	Sulfuric Acid (Elec. Tr.)	Ph. Ext/Storage
12"	Wet Flare (Trace)	Flare
2"	Nitrogen	Main rack
2"	Plant Air	"
2"	Instr. Air	"
2"	Raw Water (Elec. Tr.)	"
3"	LP Steam	"
2"	M.P. Steam	"
6"	H.P. Steam	"
4"	Stm. Cond.	"
2"	BFW	"
12"	C. W. Supply & Return	"
1 1/2"	Wash Water	Treatment/Ph. Ext.
1 1/2"	Waste Water (Elec. Tr.)	Phenosolvan/Ph. Ext.
15"	Storm Sewer (9' deep)	Storm Basin
15"	Oily Water Sewer (9' deep)	8100/Process Unit
6"	Sanitary Sewer (9' deep)	8400/Process Unit
10"	Fire Water	Ring Headers

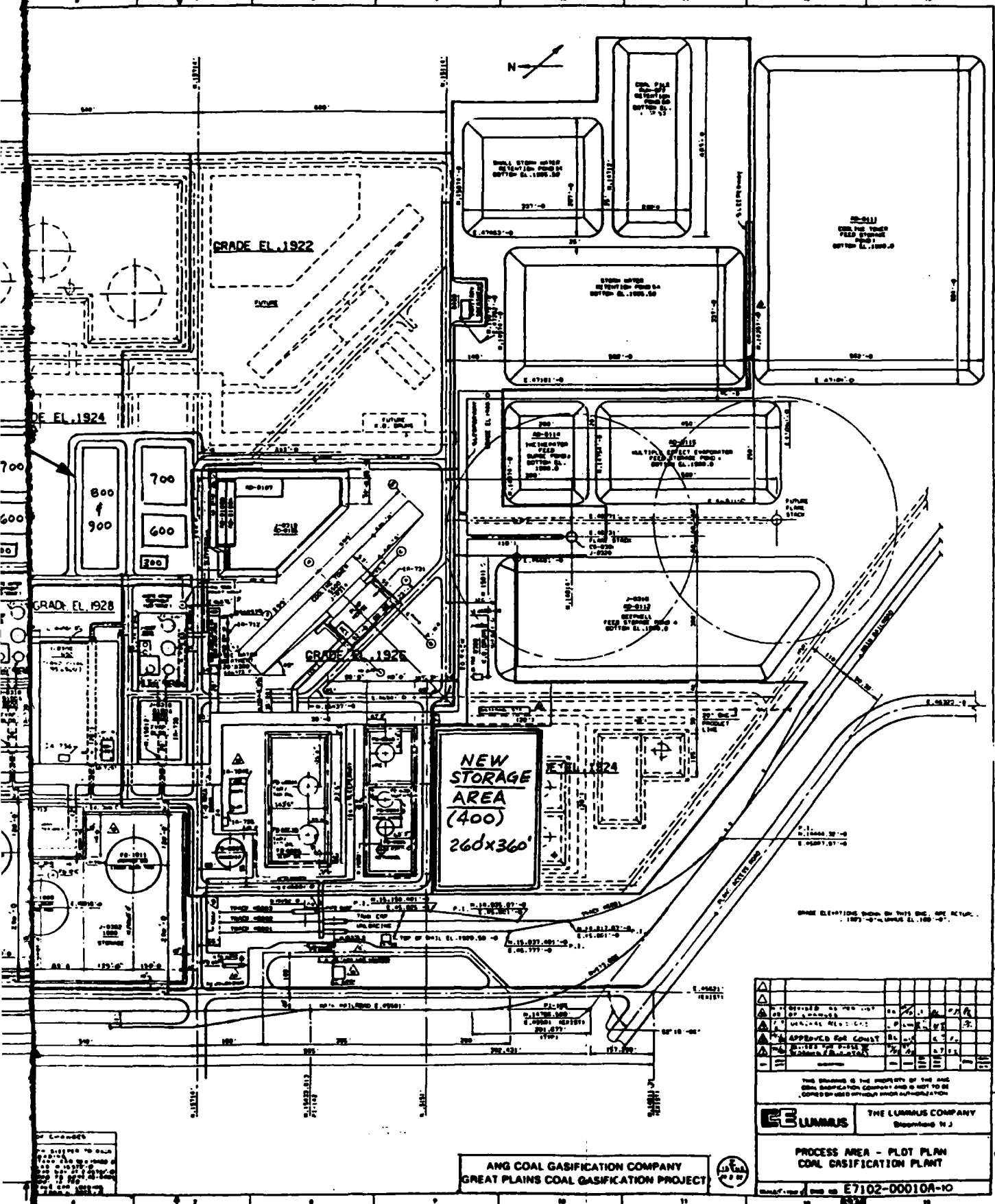
TABLE 5.1 - cont'd

INTERCONNECTING PIPING

II - NAPHTHA STREAM

<u>EST. SIZE</u>	<u>SERVICE</u>	<u>TO/FROM</u>
1 1/2"	Crude Naphtha	Storage
1 1/2"	160°F - Distillate	Storage/Dist.
1 1/2"	Blending Stock	Storage/ARU
1 1/2"	Benzene	Storage/ARU
1 1/2"	Toluene	Storage/ARU
1 1/2"	Xylene	Storage/ARU
1 1/2"	Butane	Storage
3"	Gasoline	Storage
14"	Wet Flare (Trace)	Flare
1 1/2"	Synthesis Gas	PSA/Rectisol
3"	Purge Gas	Fuel Gas/PSA & HDT
1 1/2"	Off Gas	Rectisol/HDT
2"	Nitrogen	Main Rack
2"	Plant Air	"
2"	Instr. Air	"
2"	Raw Water (Elec. Tr.)	"
6"	M.P. Steam	"
6"	H.P. Steam	"
6"	Stm Cond.	"
10"	C. W. Supply & Return	"
2"	Waste Water	Phosam/HDT
4"	Fuel Oil (Elec. Tr.)	Exist TKS/New TKS.
15"	Storm Sewer (9' deep)	Storm Basin
15"	Oily Water Sewer (9' deep)	8100/Process Unit
6"	Sanitary Sewer (9' deep)	8400/Process Unit
10"	Fire Water	Ring Headers





**COAL PREPARATION
AND HANDLING
AREA-2000**

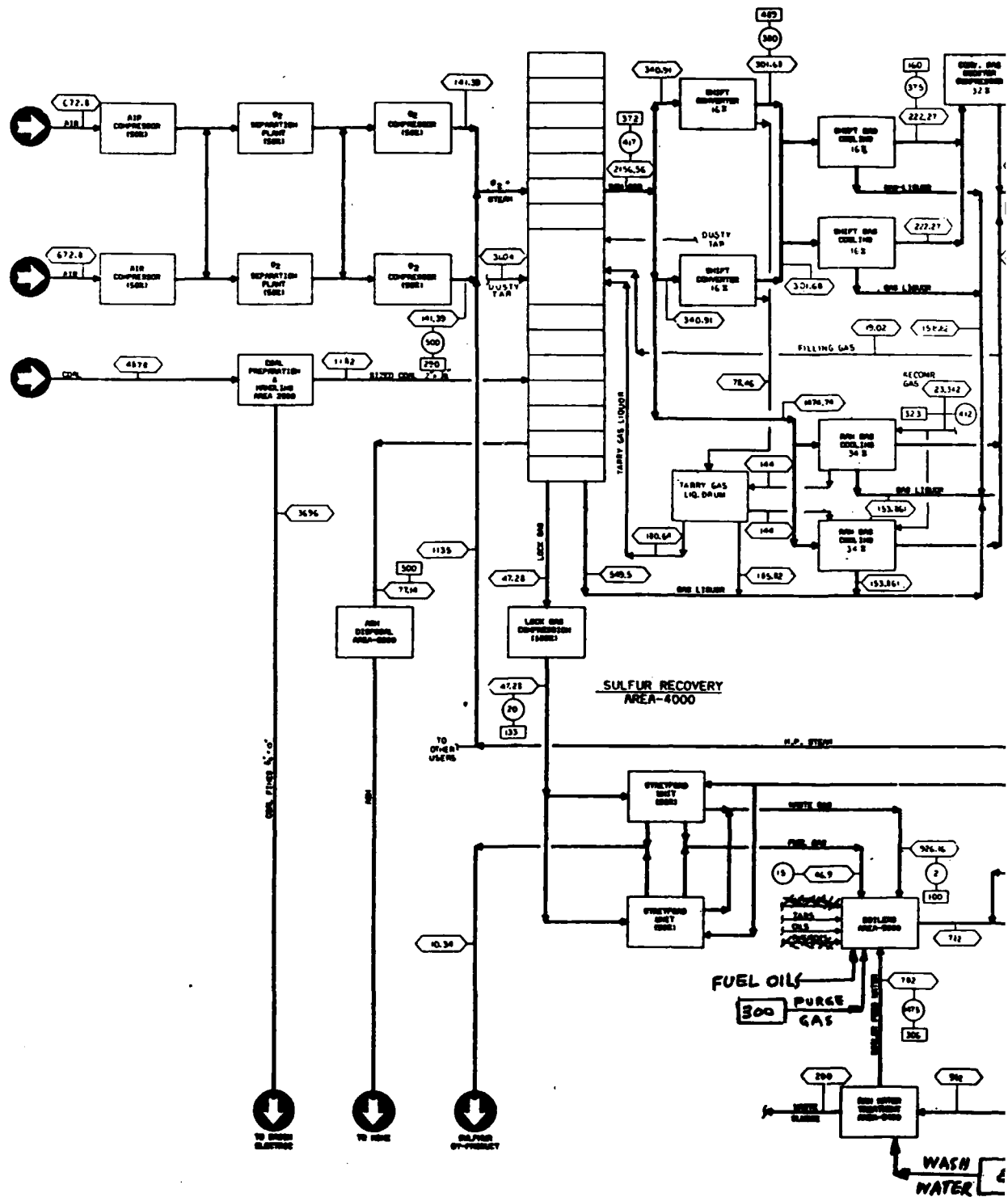
**OXYGEN PLANTS
AREA-3000**

**GASIFICATION
AREA-1100**

**SHIFT CONVERSION
AREA-1200**

**GAS COOLING
AREA-1300**

OFF GA



CASE 7

APPENDIX A
AREA 800/900
PAGE 1

CRUDE PHENOLS PRODUCTION		113 MILLION LB/YR		CRUDE PHENOL COLUMN						
STREAM FACTOR				1		13				
ON-STREAM NR/YR 7796.4										
STREAM NO										
STREAM DESCRIPTION		M.P. DEG C	B.P. DEG C	CRUDE PHENOL #/HR	PHENOL RECYCLE #/HR	o-CRESOL PHENOL #/HR	FEED #/HR	TOTAL OVERHEAD #/HR	OVERHEAD PHENOL #/HR	WATER + LIGHTS #/HR
COMPONENT										
WATER	N2O		100	724	0		724	724	55	670
METHANOL				0	0					
HEXANE				0	0					
SURFIC ACID	H2SO4			0	0					
LIGHTS				290	1		291	291		291
PYRIDINES	C5H5N		115	145	0		145	145	1	144
PHENOL	C6H5OH	43	181	4925	490	703	6118	5479	5426	53
NEUTRAL OIL				579	1		580	82	82	
o-CRESOL	CH3C6H4OH 1,2	30	191	869	0	7	877	0		
p-CRESOL	CH3C6H4OH 1,4	35.5	202	0	0	0	0	0		
m-CRESOL	CH3C6H4OH 1,3	11	202	1977	0		1978	0		
GUAIACOL	HOC6H4OCH3 1,2	32	205	0	0		0	0		
o-ETHYL PHENOL	HOC6H4C2H5		207	58	0		58	0		
2,4-XYLENOL	HOC6H3(CH3)2	26	211	1050	0		1050	0		
2,5-XYLENOL	HOC6H3(CH3)2	75	212	0	0		0	0		
2,6-XYLENOL	HOC6H3(CH3)2		212	0	0		0	0		
p-ETHYL PHENOL	HOC6H4C2H5		214	0	0		0	0		
2,3-XYLENOL	HOC6H3(CH3)2		218	0	0		0	0		
3,5-XYLENOL	HOC6H3(CH3)2	68	219	0	0		0	0		
m-ETHYL PHENOL	HOC6H4C2H5		219	109	0		109	0		
3,4-XYLENOL	HOC6H3(CH3)2	62.5	225	0	0		0	0		
CATECHOL	C6H4(OH)2 1,2	105	245	0	0		0	0		
RESIDUE				3767	0		3767	0		
RESORCINOL	C6H4(OH)2 1,3	110	281	0	0		0	0		
UNKNOWN				0	0		0	0		
TOTAL #/HR				14494	492	710	15697	6721	5564	1157
API				1.7	1.2	1.2	1.7	3.3	1.5	12.4
S.G.				1.062	1.066	1.066	1.062	1.05	1.064	0.983
#/GAL				8.857	8.890	8.890	8.858	8.756	8.873	8.197
BSD				935	32	46	1013	439	358	81
BCD				832	28	41	901	390	319	72
WT% / RECOVERY				WT %			WT %	WT %	WT %	
WATER				5.00				100	1	17.68
METHANOL										
HEXANE										
SURFIC ACID										
LIGHTS				2.00				100		
PYRIDINES				1.00				100	0.02	
PHENOL				34.00				89.55		1.40
NEUTRAL OIL				4.00				1.50	1.5	
o-CRESOL				6.00						
p-CRESOL				0.00						
m-CRESOL				13.65						
GUAIACOL										
o-ETHYL PHENOL				0.40						
2,4-XYLENOL				7.25						
2,5-XYLENOL										
2,6-XYLENOL										
p-ETHYL PHENOL										
2,3-XYLENOL										
3,5-XYLENOL										
m-ETHYL PHENOL				0.75						
3,4-XYLENOL										
CATECHOL										
RESIDUE				26.00						
RESORCINOL										
UNKNOWN										

STREAM NO	LIGHT ENDS COLUMN					PHENOL COLUMN					
	2					5					
	OVERHEAD WATER #/HR	OVERHEAD LIGHTS #/HR	BOTTOMS #/HR	FEED #/HR	WATER #/HR	OVERHEAD LIGHTS #/HR	OVERHEAD WATER #/HR	BOTTOMS #/HR	TOTAL OVERHEAD PHENOL #/HR	BOTTOMS #/HR	PHENOL PRODUCT #/HR
WATER	670		0	55	4022	0	4077	0			0
METHANOL				0		0	0	0			0
HEXANE				0		0	0	0			0
SURFUIC ACID				0		0	0	0			0
LIGHTS		291	0	291		280	0	11	11		10
PYRIDINES		144	0	145		145	0	0			0
PHENOL	18	35	639	5461	318	0	322	5456	5386	70	4896
NEUTRAL OIL			498	82		1	0	81	11	70	10
o-CRESOL			877	0		0	0	0	5		5
p-CRESOL			0	0		0	0	0	0		0
m-CRESOL			1978	0		0	0	0	5		5
GUAIACOL			0	0		0	0	0	0		0
o-ETHYL PHENOL			58	0		0	0	0	0		0
2,4-XYLENOL			1050	0		0	0	0	0		0
2,5-XYLENOL			0	0		0	0	0	0		0
2,6-XYLENOL			0	0		0	0	0	0		0
p-ETHYL PHENOL			0	0		0	0	0	0		0
2,3-XYLENOL			0	0		0	0	0	0		0
3,5-XYLENOL			0	0		0	0	0	0		0
m-ETHYL PHENOL			109	0		0	0	0	0		0
3,4-XYLENOL			0	0		0	0	0	0		0
CATECHOL			0	0		0	0	0	0		0
RESIDUE			3767	0		0	0	0	0		0
RESORCINOL			0	0		0	0	0	0		0
UNKNOWN			0	0		0	0	0	0		0
TOTAL #/HR	687	469	8976	6033	4340	425	4399	5549	5419	141	4926
API	10.0	15.1	5.9	2.6	10.0	14.4	10.0	2.0	1.2	10.0	1.2
S.G.	1	0.965	1.03	1.055	1.000	0.970	1.000	1.060	1.066	1.000	1.066
#/GAL	8.339	8.047	8.589	8.802	8.343	8.093	8.343	8.843	8.894	8.343	8.89
BCD	47	33	597	392	297	30	301	359	348	10	316
BCD	42	30	531	349	265	27	268	319	310	9	282
WT% / RECOVERY					WT %	WT %	WT %	WT %	WT %	WT %	WT %
WATER					66.67		100				91
METHANOL											91
HEXANE											91
SURFUIC ACID											91
LIGHTS								0.2	0.2		91
PYRIDINES						100	0				91
PHENOL					8		8	99.8	99.4	50	91
NEUTRAL OIL						1	0		0.2	50	91
o-CRESOL									0.1		91
p-CRESOL											91
m-CRESOL											91
GUAIACOL									0.1		91
o-ETHYL PHENOL											91
2,4-XYLENOL											91
2,5-XYLENOL											91
2,6-XYLENOL											91
p-ETHYL PHENOL											91
2,3-XYLENOL											91
3,5-XYLENOL											91
m-ETHYL PHENOL											91
3,4-XYLENOL											91
CATECHOL											91
RESIDUE											91
RESORCINOL											91
UNKNOWN											91

STREAM NO	ACID FLASH				TAR WASHING				EXTRACTOR COLUMN		
	7		4		8						
	PHENOL RECYCLE	FEED	SULFURIC ACID	TOTAL FEED	OVERHEAD	TAR OIL	WASH WATER	WASTE WATER	WASHED TAR	FEED FR ACID FLASH	HEXANE SOLVENT
COMPONENT	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
WATER	0	0	6	6	6	0	1246	1246	0	6	0
METHANOL	0				0				0	0	0
HEXANE	0				0				0	0	11372
SURFUIIC ACID	0		285	285	0	285		285	0	0	0
LIGHTS	1	0		0	0	0			0	0	0
PYRIDINES	0	0		0	0	0			0	0	0
PHENOL	490	710		710	710	0			0	710	0
NEUTRAL OIL	1	569		569	569	0			0	569	0
o-CRESOL	0	877		877	877	0			0	877	0
p-CRESOL	0	0		0	0	0			0	0	0
m-CRESOL	0	1978		1978	1938	40			40	1938	0
GUAIACOL	0	0		0	0	0			0	0	0
o-ETHYL PHENOL	0	58		58	57	1			1	57	0
2,4-XYLENOL	0	1050		1050	1029	21			21	1029	0
2,5-XYLENOL	0	0		0	0	0			0	0	0
2,6-XYLENOL	0	0		0	0	0			0	0	0
p-ETHYL PHENOL	0	0		0	0	0			0	0	0
2,3-XYLENOL	0	0		0	0	0			0	0	0
3,5-XYLENOL	0	0		0	0	0			0	0	0
m-ETHYL PHENOL	0	109		109	106	2			2	106	0
3,4-XYLENOL	0	0		0	0	0			0	0	0
CATECHOL	0	0		0	0	0			0	0	0
RESIDUE	0	3767		3767	377	3390			3390	377	0
RESORCINOL	0	0		0	0	0			0	0	0
UNKNOWN	0	0		0	0	0			0	0	0
TOTAL #/HR	493	9117	291	9408	5669	3739	1246	1532	3454	5669	11372
API	1.2	6.0		4.1	11.9					11.9	
S.G.	1.066	1.029	1.83	1.043	0.987	1.140	1.000	1.100	1.100	0.987	0.660
#/GAL	8.89	8.585	15.27	8.703	8.23	9.51	8.34	9.18	9.18	8.23	5.51
BSD	32	607	11	618	393	225	85	95	215	393	1180
BCD	28	540	10	550	350	200	76	85	191	350	1050
WT% / RECOVERY	WT %		WT %		WT %		WT%				WT %
WATER	9		2		100		33.33				
METHANOL	9										
HEXANE	9										100
SURFUIIC ACID	9		3								
LIGHTS	9				100						
PYRIDINES	9				100						
PHENOL	9				100						
NEUTRAL OIL	9				100						
o-CRESOL	9				100						
p-CRESOL	9				98						
m-CRESOL	9				98						
GUAIACOL	9				98						
o-ETHYL PHENOL	9				98						
2,4-XYLENOL	9				98						
2,5-XYLENOL	9				98						
2,6-XYLENOL	9				98						
p-ETHYL PHENOL	9				98						
2,3-XYLENOL	9				98						
3,5-XYLENOL	9				98						
m-ETHYL PHENOL	9				98						
3,4-XYLENOL	9				98						
CATECHOL	9				95						
RESIDUE	9				10						
RESORCINOL	9				5						
UNKNOWN	9				5						

STREAM NO	HEXANE COLUMN					METHANOL COLUMN		DRYING COLUMN	o-CRESOL COLUMN		
	10							11	13	12	
	METHANOL/ WATER SOLVENT #/HR	OVERHEAD #/HR	BOTTOMS #/HR	HEXANE OVERHEAD #/HR	NEUTRAL OIL BOTTOMS #/HR	METHANOL/ WATER OVERHEAD #/HR	NET CRESYLIC ACID #/HR	DRYING COLUMN OVHD #/HR	DRY CRESYLIC ACID #/HR	o-CRESOL PHENOL #/HR	o-CRESOL #/HR
WATER	1765	0	1771	0	0	1594	177	177	0		
METHANOL	3278	0	3278	0	0	3278	0				
HEXANE	0	11372	0	11372	0	0	0				
SURFUIC ACID	0	0	0	0	0	0	0		0		
LIGHTS	0	0	0	0	0	0	0		0		
PYRIDINES	0	0	0	0	0	0	0		0		
PHENOL	0	0	710	0	0	0	710		710	703	7
NEUTRAL OIL	0	569	0	0	569	0	0		0		
o-CRESOL	0	0	877	0	0	0	877		877	7	826
p-CRESOL	0	0	0	0	0	0	0		0	0	0
m-CRESOL	0	0	1938	0	0	0	1938		1938		17
GUAIACOL	0	0	0	0	0	0	0		0		
o-ETHYL PHENOL	0	0	57	0	0	0	57		57		
2,4-XYLENOL	0	0	1029	0	0	0	1029		1029		
2,5-XYLENOL	0	0	0	0	0	0	0		0		
2,6-XYLENOL	0	0	0	0	0	0	0		0		
p-ETHYL PHENOL	0	0	0	0	0	0	0		0		
2,3-XYLENOL	0	0	0	0	0	0	0		0		
3,5-XYLENOL	0	0	0	0	0	0	0		0		
m-ETHYL PHENOL	0	0	106	0	0	0	106		106		
3,4-XYLENOL	0	0	0	0	0	0	0		0		
CATECHOL	0	0	0	0	0	0	0		0		
RESIDUE	0	0	377	0	0	0	377		377		
RESORCINOL	0	0	0	0	0	0	0		0		
UNKNOWN	0	0	0	0	0	0	0		0		
TOTAL #/HR	5043	11940	10142	11372	569	4871	5271	177	5094	710	849
API							8.6	8.6	8.6	2	5.9
S.G.	0.878	0.669	0.950	0.660	0.914	0.876	1.030	1.000	1.030	1.066	1.035
#/GAL	7.33	5.58	7.93	5.51	7.63	7.31	8.59	8.34	8.59	8.89	8.63
BSD	393	1222	731	1180	43	381	351	12	339	46	56
BCD	350	1088	651	1050	38	339	312	11	301	41	50
WTX / RECOVERY	WT %	RECOVERY	RECOVERY						RECOVERY	RECOVERY	
WATER	35		100			90	10				
METHANOL	65		100			100					
HEXANE		100		100							
SURFUIC ACID											
LIGHTS											
PYRIDINES											
PHENOL			100							99	1
NEUTRAL OIL		100									
o-CRESOL			100							1	95
p-CRESOL											
m-CRESOL											2
GUAIACOL											
o-ETHYL PHENOL											
2,4-XYLENOL											
2,5-XYLENOL											
2,6-XYLENOL											
p-ETHYL PHENOL											
2,3-XYLENOL											
3,5-XYLENOL											
m-ETHYL PHENOL											
3,4-XYLENOL											
CATECHOL											
RESIDUE											
RESORCINOL											
UNKNOWN											

STREAM NO	m,p-CRESOL COLUMN			XYLENOL COLUMN				
	14		o-ETHYL PHENOL	15		16	6	
	o-CRESOL COLUMN BOTTOMS	m/pCRESOL		o-CRESOL COLUMN BOTTOMS	XYLENOLS	HEAVIES	PHEN EXT NEUTRAL OIL	TOTAL NEUTRAL OIL
COMPONENT	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR	#/HR
WATER	0	0		0	0	0	0	0
METHANOL							0	0
HEXANE							0	0
SURFUIC ACID	0	0		0	0	0	0	0
LIGHTS	0	0		0	0	0	280	280
PYRIDINES	0	0		0	0	0	145	145
PHENOL	0	0		0	0	0	0	0
NEUTRAL OIL	0	0		0	0	0	570	570
o-CRESOL	44	44		0	0	0	0	0
p-CRESOL	0	0		0	0	0	0	0
m-CRESOL	1922	1826	0	96	96	0	40	40
GUAIACOL	0			0	0	0	0	0
o-ETHYL PHENOL	57	54	0	3	3	0	1	1
2,4-XYLENOL	1029	51		978	929	49	21	70
2,5-XYLENOL	0			0		0	0	0
2,6-XYLENOL	0			0		0	0	0
p-ETHYL PHENOL	0			0		0	0	0
2,3-XYLENOL	0			0		0	0	0
3,5-XYLENOL	0			0		0	0	0
m-ETHYL PHENOL	106			106	49	58	2	60
3,4-XYLENOL	0			0		0	0	0
CATECHOL	0			0		0	0	0
RESIDUE	377			377	0	377	3390	3767
RESORCINOL	0			0		0	0	0
UNKNOWN	0			0		0	0	0
TOTAL #/HR	3535	1975	0	1560	1077	483	4448	4931
API	5.9	5.9	5.9	14.4	13.6	-0.5	2.1	0.7
S.G.	1.024	1.034	1.030	1.010	0.975	1.08	1.059	1.07
#/GAL	8.54	8.63	8.59	8.43	8.13	9.01	8.834	8.93
BSD	237	131	0	106	76	31	288	316
BCD	210	116	0	94	67	27	256	281

MTX / RECOVERY RECOVERY RECOVERY RECOVERY

WATER
METHANOL
HEXANE
SURFUIC ACID
LIGHTS
PYRIDINES
PHENOL
NEUTRAL OIL
o-CRESOL
p-CRESOL
m-CRESOL
GUAIACOL
o-ETHYL PHENOL
2,4-XYLENOL
2,5-XYLENOL
2,6-XYLENOL
p-ETHYL PHENOL
2,3-XYLENOL
3,5-XYLENOL
m-ETHYL PHENOL
3,4-XYLENOL
CATECHOL
RESIDUE
RESORCINOL
UNKNOWN

95

95

0

5

95

5

0

REFINERY PROCESSOR PROPERTIES SET

STREAM ID.	5	5	5	7	8	9	10	11
STREAM NAME	MAKE-UP GAS	REACTOR INLE	REACTOR FFEL	WASH WATER	COLD SEP LIQ	PURGE GAS	RECYCLE GAS	
STREAM PHASE	VAPOR	VAPOR	VAPOR	LIQUID	LIQUID	VAPOR	VAPOR	
TEMPERATURE, DEG F	70.0000	425.0000	475.0000	190.0000	120.0000	120.0000	120.0000	
PRESSURE, PSIA	340.0000	400.0000	775.0000	735.0000	710.0000	710.0000	710.0000	
RATE, LB MOLS/HR	31.0000	277.6176	264.3793	138.7732	82.3024	5.3009	168.4035	
RATE, LB /HR	62.5395	7324.1631	7321.4014	2499.9995	6446.4883	21.7711	691.6453	
ENTHALPY MM BTU /HR	-0.0644	2.2362	2.5851	0.1750	0.2584	-0.0062	-0.1986	
ENTHALPY BTU /LB	-1030.3195	303.3162	353.9320	69.9971	40.0906	-286.8285	-286.8285	
MOLECULAR WEIGHT....	2.0174	26.3922	27.6928	18.0150	78.3268	4.1071	4.1071	
*** VAPOR PHASE ***								
RATE, LB /HR	62.5395	7324.1631	7321.4014	0.0000	0.0000	21.7711	691.6453	
ACT-RATE, FT3/SEC	0.15	0.89	0.93	0.00	0.00	0.01	0.42	
STD-RATE MM FT3/DAY	0.29	2.53	2.41	0.00	0.00	0.05	1.53	
CP, BTU /LB F	3.4201	0.6139	0.6083	0.0000	0.0000	1.8091	1.8091	
MOLECULAR WEIGHT....	2.0174	26.3922	27.6928	0.0000	0.0000	4.1071	4.1071	
ACT-DENS, LB /FT3	0.1189	2.2787	2.1994	0.0000	0.0000	0.4369	0.4369	
COMPRESSIBILITY (Z)	1.0145	0.9757	0.9773	0.0000	0.0000	1.0260	1.0260	
*** LIQUID PHASE ***								
RATE, LB /HR	0.0000	0.0000	0.0000	2499.9995	6446.4883	0.0000	0.0000	
ACT-RATE, BBL/DAY	0.00	0.00	0.00	172.41	541.70	0.00	0.00	
STD. LV RATE BBL/HR	0.00	0.00	0.00	7.15	21.83	0.00	0.00	
CP, BTU /LB F	0.0000	0.0000	0.0000	0.9941	0.4159	0.0000	0.0000	
MOLECULAR WEIGHT....	0.0000	0.0000	0.0000	18.0150	78.3268	0.0000	0.0000	
ACT-DENS, LB /FT3	0.0000	0.0000	0.0000	61.9844	50.8694	0.0000	0.0000	
STD. API GRAVITY....	0.0000	0.0000	0.0000	10.0635	34.1255	0.0000	0.0000	
*** DRY BASIS ***								
RATE, LB /HR	0.0000	0.0000	0.0000	0.0000	6443.5264	0.0000	0.0000	
ACT-RATE, FT3/SEC	0.0000	0.0000	0.0000	0.0000	78.4475	0.0000	0.0000	
STD-RATE MM FT3/DAY	0.00	0.00	0.00	0.00	10.1772	0.0000	0.0000	
UOP K	0.0000	0.0000	0.0000	0.0000	-75.8712	0.0000	0.0000	
FLASH POINT, DEG F	0.0000	0.0000	0.0000	0.0000	519.0099	0.0000	0.0000	
CRIT. TEMP, F	0.0000	0.0000	0.0000	0.0000	678.9440	0.0000	0.0000	
CRIT. PRES, PSIA	0.0000	0.0000	0.0000	0.0000				
*** VAPOR PHASE ***								
RATE, LB /HR	62.5395	7316.9846	7187.2246	0.0000	0.0000	21.5452	684.4674	
ACT-RATE, FT3/SEC	0.15	0.89	0.90	0.00	0.00	0.01	0.42	
STD-RATE MM FT3/DAY	0.01	0.11	0.10	0.00	0.00	0.00	0.06	
CP, BTU /LB F	3.4201	0.6139	0.6113	0.0000	0.0000	1.8233	1.8233	
MOLECULAR WEIGHT....	2.0174	26.3942	27.2733	0.0000	0.0000	4.0741	4.0741	
ACT-DENS, LB /FT3	0.1189	2.2793	2.2128	0.0000	0.0000	0.4532	0.4532	
COMPRESSIBILITY (Z)	1.0145	0.9756	0.9768	0.0000	0.0000	1.0261	1.0261	
VISCOSITY, CP	0.0078	0.0185	0.0191	0.0000	0.0000	0.0106	0.0106	
*** LIQUID PHASE ***								
RATE, LB /HR	0.0000	0.0000	0.0000	0.0000	6443.5264	0.0000	0.0000	
ACT-RATE, BBL/DAY	0.00	0.00	0.00	0.00	541.49	0.00	0.00	
STD. LV RATE BBL/HR	0.0000	0.0000	0.0000	0.0000	7.15	0.0000	0.0000	
UOP K	0.0000	0.0000	0.0000	0.0000	-75.8712	0.0000	0.0000	
FLASH POINT, DEG F	0.0000	0.0000	0.0000	0.0000	519.0099	0.0000	0.0000	
CRIT. TEMP, F	0.0000	0.0000	0.0000	0.0000	678.9440	0.0000	0.0000	
CRIT. PRES, PSIA	0.0000	0.0000	0.0000	0.0000				

NAPHTHA HDY PRODUCT STRIPPER

STREAM ID.	12	13	15	17
STREAM NAME.....	SOUR H2O	STAB PRO	STAB OFFGAS	STAB SOUR H2O
STREAM PHASE.....	LIQUID	LIQUID	VAPOR	LIQUID
TEMPERATURE, DEG F	120.0000	287.5287	100.0000	100.0000
PRESSURE, PSIA	710.0000	70.0000	60.0000	60.0000
RATE LB MOLS/HR	147.1456	73.8262	8.4432	0.0309
RATE LB /HR	2661.4951	6135.1143	313.3105	0.5570
ENTHALPY MM BTU /HR	0.0780	0.6899	0.0453	0.0000
ENTHALPY BTU /LB	29.3215	112.4480	208.5772	67.9931
MOLECULAR WEIGHT....	18.0875	83.0998	37.1079	18.0150
*** VAPOR PHASE ***				
RATE LB /HR	0.0000	0.0000	313.3105	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.23	0.00
STD.RATE MM FT3/DAY	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.3973	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	37.1079	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.3844	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9645	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	2661.4951	6135.1143	0.0000	0.5570
ACT.RATE BBL/DAY	185.82	576.27	0.00	0.04
STD. LV RATE BBL/HR	7.66	20.33	0.00	0.00
CP, BTU /LB F	1.1616	0.4890	0.0000	0.9977
MOLECULAR WEIGHT....	18.0875	83.0998	0.0000	18.0150
ACT.DENS LB /FT3	61.2244	45.5081	0.0000	61.9843
STD. API GRAVITY....	10.8946	32.5122	0.0000	10.0635
*** DRY BASIS ***				
RATE LB /HR	37.6836	6135.1016	0.0000	0.0000
MOLECULAR WEIGHT....	25.1254	83.1004	0.0000	0.0000
UOP K	10.2499	10.0651	0.0000	0.0000
FLASH POINT, DEG F	-5.3046	0.7399	0.0000	0.0000
CRIT. TEMP, PSIA	216.9373	561.6288	0.0000	0.0000
CRIT. PRES, PSIA	1412.3528	667.3295	0.0000	0.0000
*** VAPOR PHASE ***				
RATE LB /HR	0.0000	0.0000	310.9178	0.0000
ACT.RATE FT3/SEC	0.00	0.00	0.22	0.00
STD.RATE MM FT3/DAY	0.00	0.00	0.00	0.00
CP, BTU /LB F	0.0000	0.0000	0.3971	0.0000
MOLECULAR WEIGHT....	0.0000	0.0000	37.4130	0.0000
ACT.DENS LB /FT3	0.0000	0.0000	0.3877	0.0000
COMPRESSIBILITY (Z).	0.0000	0.0000	0.9640	0.0000
VISCOSITY, CP	0.0000	0.0000	0.0102	0.0000
*** LIQUID PHASE ***				
RATE LB /HR	37.6836	6135.1016	0.0000	0.0000
ACT.RATE BBL/DAY	4.07	576.27	0.00	0.00
CP, BTU /LB F	0.8241	0.4390	0.0000	0.0000
MOLECULAR WEIGHT....	25.1254	83.1004	0.0000	0.0000
ACT.DENS LB /FT3	39.6012	45.5080	0.0000	0.0000
STD. API GRAVITY....	61.7618	32.5122	0.0000	0.0000