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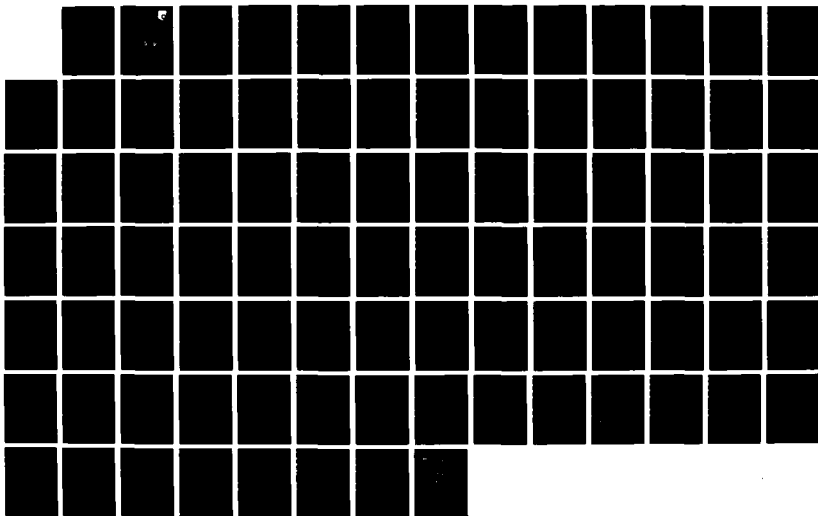
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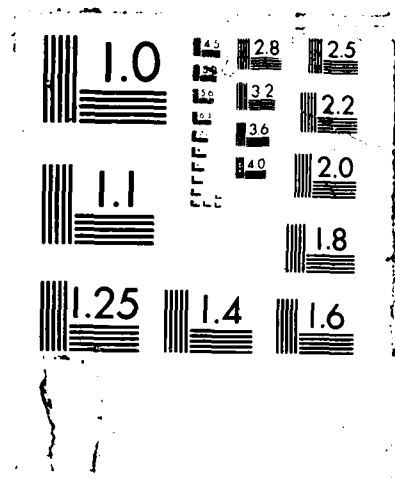
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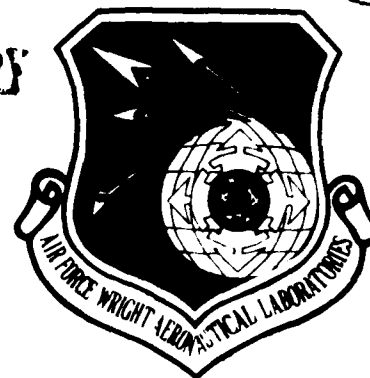




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Part III

AVIATION TURBINE FUELS FROM TAR SANDS
BITUMEN AND HEAVY OILS

Part III Laboratory Sample Production

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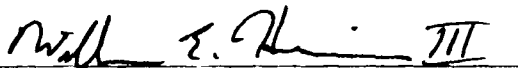
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This technical report has been reviewed and is approved for publication.



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19. ABSTRACT (Continue on reverse if necessary and identify by block number) Samples of specification JP-4 Mil-T-5624L, JP-8 Mil-T-83133A, and variable quality JP-4 samples were produced via pilot plant operations. Data generated from Phases I, II, and III, were used to 1) optimize the processing scheme, 2) generate process material and energy balances for a commercial sized plant, and 3) provide a detailed final flow diagram of the processing scheme. A final economic analysis was performed based on all contract data available.					
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FOREWORD

This project was sponsored by the U.S. Air Force Wright Aeronautical Laboratories (AFWL) Air Force Systems Command, under Contract No. F33615-83-C-2301. The work herein was performed during the period July 1, 1983 to September 30, 1986. This interim report describes the Phase III efforts of Ashland Petroleum Company Research and Development personnel in the Pilot Plant preparation of JP-4 and JP-8 samples and in the computer modeling optimization study of the overall process.

The authors wish to acknowledge the contributions of the following individuals: Dr. M. M. Mitchell, Jr., Vice President and Director of Research and Development; Mr. Robert E. Stone, Computer Aided Evaluation Engineer; and Ms. Sue White, Correspondence Word Processor. The helpful suggestions of the Air Force Contract Project Engineer, Ms. Teresa Planeaux, were greatly appreciated and were of benefit throughout the contract program.

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SECTION I

INTRODUCTION

The traditional source of aviation fuels has been the refining of petroleum. In recent years, the consumption of petroleum products in the United States has exceeded our country's discovery and development of new oil production. The lessening world supply of crude oil, the increased cost of this crude oil, and specifically the dependence of the United States on foreign oil sources were vividly demonstrated during the Arab oil embargo in 1973, as well as the 1979 Iranian crisis. All of these conditions served to emphasize the need for the development of new energy sources within the United States to ensure a continued national energy supply. While recent trends show adequate supply and lowered cost, a secure and reliable supply of military fuels is still essential for our national defense. For this reason, the Department of Defense and the Department of Energy have set into motion programs for the development of fuels from tar sand and heavy oil deposits located in the United States.

Contract No. F33615-83-C-2301
The Research and Development Department of Ashland Petroleum Company has been awarded Contract No. F33615-83-C-2301 to provide sample quantities of aviation turbine fuel derived from tar sands and heavy oil feedstocks for testing and evaluation in programs sponsored by the Air Force Wright Aeronautical Laboratories (AFWAL). The goals to be pursued

under this program were (1) provide samples of variable quality military fuels which can be economically produced from tar sands and heavy oils by methods which shall be disclosed to the Air Force; (2) develop a model of the processing method to project economic data based on throughputs which minimize product costs and maximize overall plant thermal efficiency; (3) provide a minimum overall efficiency of 70 percent, based on crude charge, product yield and utility consumptions, including the hydrogen consumption; and, (4) produce a full slate of military transportation fuels. This slate of fuels was to include motor gasoline, aviation turbine fuels (grades JP-4 and JP-8), and residual fuel products. The yields of residual fuel were limited to no more than 10 percent of the product slate while maximizing the yield of aviation turbine fuel, grade JP-4 or JP-8.

This program was divided into three phases. Phase I commenced on July 1, 1983 and was completed on June 15, 1984, with the primary objectives of evaluating the U. S. tar sand/heavy oil resource base and performing a preliminary process analysis. Conceptual flow diagrams, yields, and process economics were developed which demonstrated the potential of this process.¹ Phase II was initiated on April 2, 1984 and was completed on January 31, 1985. This phase consisted of two major tasks: (1) an evaluation of operating condition impacts on process performance, and (2) production of small (500 milliliter) samples of variable quality aviation turbine fuels.^{2,3} Phase II evaluated two heavy oils, (Hondo, San Ardo) and two bitumen

(Westken, Sunnyside) feedstocks. Phase III was initiated on February 1, 1985 and was completed in July 1, 1986, with the objective of producing larger scale samples of military fuels. Samples were provided of conventional specification JP-4 and JP-8, variable quality JP-4, plus gasoline and residual fuel components. An overall economic optimization via computer modeling was completed as required, and analysis of all program data were performed during Phase III. This document summarizes and reports these efforts.

Process Description

The process selected for primary evaluation is Ashland's Reduced Crude Conversion (RCCSM) process technology. This process has been developed based on laboratory, demonstration, and commercial scale equipment. A 40,000 BPD RCCSM unit has been successfully operated at Catlettsburg, Kentucky, since April 1983. A companion ARTSM Asphalt Residuum Treatment (ARTSM) unit is also in use at Catlettsburg. Details of each of these processes, and recent commercial experience, have been published elsewhere.⁴ Adaptations of these technologies were developed under this

ARTSM is a service mark of Engelhard Corporation for professional services relating to selective vaporization processes for removing contaminants from petroleum feedstocks.

RCCSM is a registered service mark of Ashland Oil, Inc., for technical assistance and consulting services in connection with processes for heavy oil cracking and related catalysts.

program which allowed processing of bitumen stocks. The overall process flow sheet for this study is shown in Figure 1.

Feedstock

The Phase III feedstock selected by the Air Force for this program was Westken bitumen. This material was produced by the Kensyntar project from a deposit located in Edmonson County, Kentucky, near the southeastern rim of the Illinois basin. The Westken bitumen has a 10.4° API gravity, a high metals content, high pour point and a significant residuum content. Distillation yields show virtually no virgin turbine fuel and about 50 volume percent heavy gas oil (600-1000°F). The hydrogen content of this feed is low compared to conventional crude oils. Both sulfur and nitrogen are moderate, with the sulfur content lower than that of many conventional sour crudes. Salt and inorganic contaminants are a primary concern due to potential refining catalyst poisoning. This feedstock was the most difficult material evaluated in Phase II, and represented a severe processing challenge. Detailed analyses of this material are available in the Phase II report.³

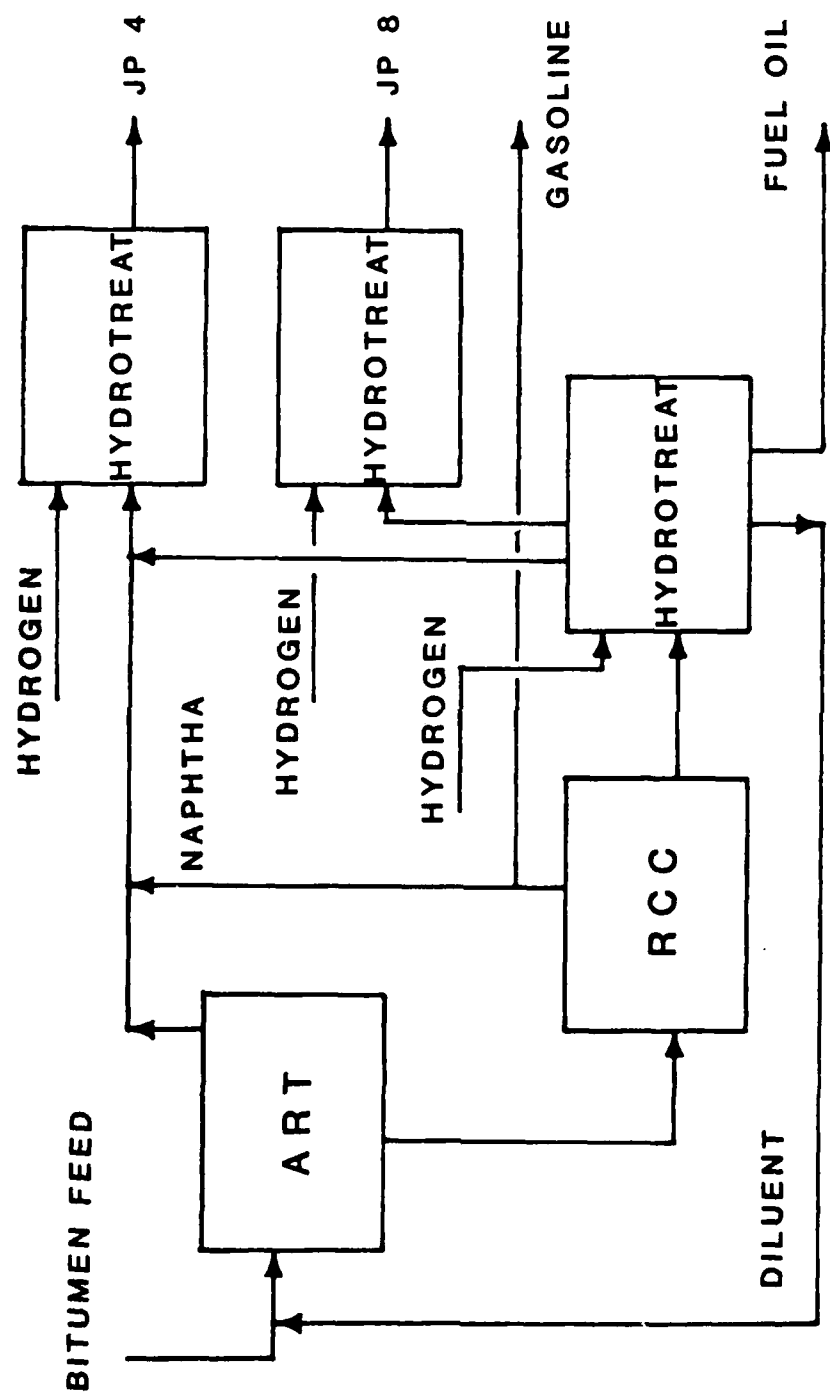


Figure 1. Simplified Process Flow

SECTION II

SAMPLE PREPARATION

Introduction

Two primary objectives were addressed in the Phase III sample preparation effort: (1) prepare five, 5 to 15 gallon samples of jet fuel, and (2) develop yield data and product analyses for input into the economic model.

Phase II results from this program revealed that a diluent was necessary to facilitate the handling of the whole Westken crude bitumen and to attain the required conversion. The Phase III effort was designed to use a Westken-derived diluent for processing to ensure purity of the final products. As a result, the initial effort in the Phase III program was to prepare a Westken-derived process diluent to simulate a recycle stream that would be used in the commercial process, followed by conversion and final fuel preparation steps.

After the diluent preparation, two complete conversion loops (Loops 1 and 2) were repeated in an effort to allow the recycle diluent properties to converge. Although in a commercial process the units would be operating simultaneously and continuously, this is not possible in the pilot plant operations because the equipment is not configured as an integrated refinery. Loops 1 and 2 were nearly identical with the primary difference being the attempt at desalting and the use of

a true process diluent stream in Loop 2. Detailed stream flows and definitions are shown in Appendix A.

Diluent Preparation

A simulated diluent was first prepared from raw Westken bitumen to ensure that all final sample materials were truly Westken derived. Since the raw bitumen was not amenable to processing as-received, the bitumen was enriched with additional Westken gas oil prepared by distillation, followed by cracking and hydrotreating as in the normal process schematic.

Diluent preparation was started with the separation of a nominal <1000°F gas oil cut made from whole crude Westken bitumen. A typical analysis of the crude bitumen is shown in Table 1; some variability was found from drum to drum. The separation was performed in a wiped film evaporator to provide a minimum residence time and relatively low temperatures to preclude thermal degradation; no projection or intent for commercial processing by this method was implied. Six drums of crude Westken were processed which produced a total of 1175 lbs. of gas oil (49.2% of feed). The 1000°F+ bottoms from this separation were discarded.

This gas oil was then mixed in a one-to-one weight ratio with crude Westken bitumen and fed to a pilot scale circulating RCCSM unit (RCR) having some of the same features as Ashland's commercial unit.

TABLE 1

DILUENT PREPARATION -
BITUMEN FRACTIONATION AND PRODUCTS

	<u>Bitumen</u>	<u>Gas Oil</u>	<u>50% Blend of Bitumen and Gas Oil</u>
Gravity, °API	10.4	19.6	15.5
Elemental Analysis, Wt%			
Sulfur	1.66	1.21	
Nitrogen	0.23	0.17	0.20
Basic Nitrogen	-	0.059	0.13
Viscosity, @ 210°F, cs	186	6.58	22.8
Pour Point, °F	65	-10	20
Ramsbottom Carbon	11.0	-	4.5
Metals, ppm:			
Nickel	63	-	33
Vanadium	229	-	98
Iron	335	-	239
Sodium	541	-	324

TABLE 2

DILUENT PREPARATION -
BLENDED GAS OIL/BITUMEN CRACKING RESULTS

	<u>Week 1</u>	<u>Week 2</u>
Catalyst:Oil Ratio	14.9	16.9
Temperature, °F	900	900
Water Injection, % Feed	12.8	10.6
Products, Wt%		
Dry Gas	2.7	3.9
Wet Gas	7.2	8.1
Gasoline	27.8	27.9
430°F+	45.6	42.9
Coke	16.7	17.3
Conversion, Wt%	54.4	57.1

The catalyst used was an equilibrium sample from the commercial RCCS^m unit, possessing good bottoms cracking ability but low to moderate activity. The 50% mixture of bitumen and gas oil was found to be difficult to process, requiring the ratio to be raised to 60/40 gas oil/bitumen. Yield patterns changed during the run (Table 2) due to the accumulation of sodium, iron, nickel, and vanadium on the catalyst. The Microactivity Test (MAT) conversion dropped from 57 to 25 volume percent and the coke factor, a relative indication of the amount of coke that would be produced, doubled. Observed coke yields increased from 12 percent at the start of run to 18 percent at the end. These effects illustrate the need for ARTS^m pretreatment of this feedstock.

The composite cracked product was distilled at 330°F. The +330°F portion was hydrotreated, and the hydrotreated product was used as the ART diluent (Table 3). Universal laboratory reactors were used, each charged with a commercial nickel-molybdate catalyst diluted 50/50 with Ottawa sand. Hydrogen consumption averaged 590 scf/bbl, typical for feedstocks of this type and hydrogenation severity. Catalyst deactivation during the run was detected by a slight decrease in API gravity of the products.

TABLE 3

DILUENT PREPARATION HYDROTREATING SUMMARYOPERATING CONDITIONS

Temperature, °F	685
Pressure, PSIG	1225
LHSV, Hr ⁻¹	2.04
Hydrogen Rate, SCFB	3096

RESULTS

Liquid Yield, Wt%	99.6
Hydrogen Consumption, SCFB	590

AVERAGE LIQUID PROPERTIES

	<u>Feed</u>	<u>Product</u>
°API	20.0	24.4
Elemental Analyses, Wt%		
Hydrogen	10.9	11.7
Sulfur	1.15	0.04
Nitrogen	0.11	0.04
Viscosity, cp @ 100°F	6.35	5.20
@ 210°F	1.80	1.63
Hydrocarbon Types, %		
Saturates	37.5	40.2
Monoaromatics	21.5	33.4
Diaromatics	14.5	8.8
>Diaromatics	20.0	13.3
Polar & Asphaltenes	6.5	4.3

LOOP 1 Conversion

Loop 1 was the first complete cycle of the process, as shown in example form in Figure 2. A mixture of Westken crude bitumen and diluent (hydrotreated Westken cycle oil from the diluent preparation loop) was processed in the ARTSM mode to remove metals and to reduce the ramsbottom carbon content, (Table 4). Equilibrium ARTCAT from the commercial unit was used for these tests. One test was made using diluent alone so that net bitumen yields could be calculated (Table 5). Products were fractionated into an I-330°F naphtha, a 330-430°F kerosene, and a 430°F+ bottoms. The naphtha was caustic washed and put in cold storage for use in blending the final sample. The 330-430°F portion was segregated for blending with RCCSM products prior to hydrogenation.

The Westken 430°F+ ARTSM bottoms were cracked in the FCR unit (a second, smaller circulating pilot cracking unit) over commercial equilibrium catalyst. Four tests were made at varying conversion levels to determine the conditions for producing maximum transportation fuels and four additional extended runs were then made to produce liquid product for subsequent diluent preparation and jet fuel blending. Two additional tests were made at maximum transportation fuel conditions on the 430°F+ diluent alone, so that bitumen yields could be calculated for use in the final process model. These results are summarized in Table 5. The maximum

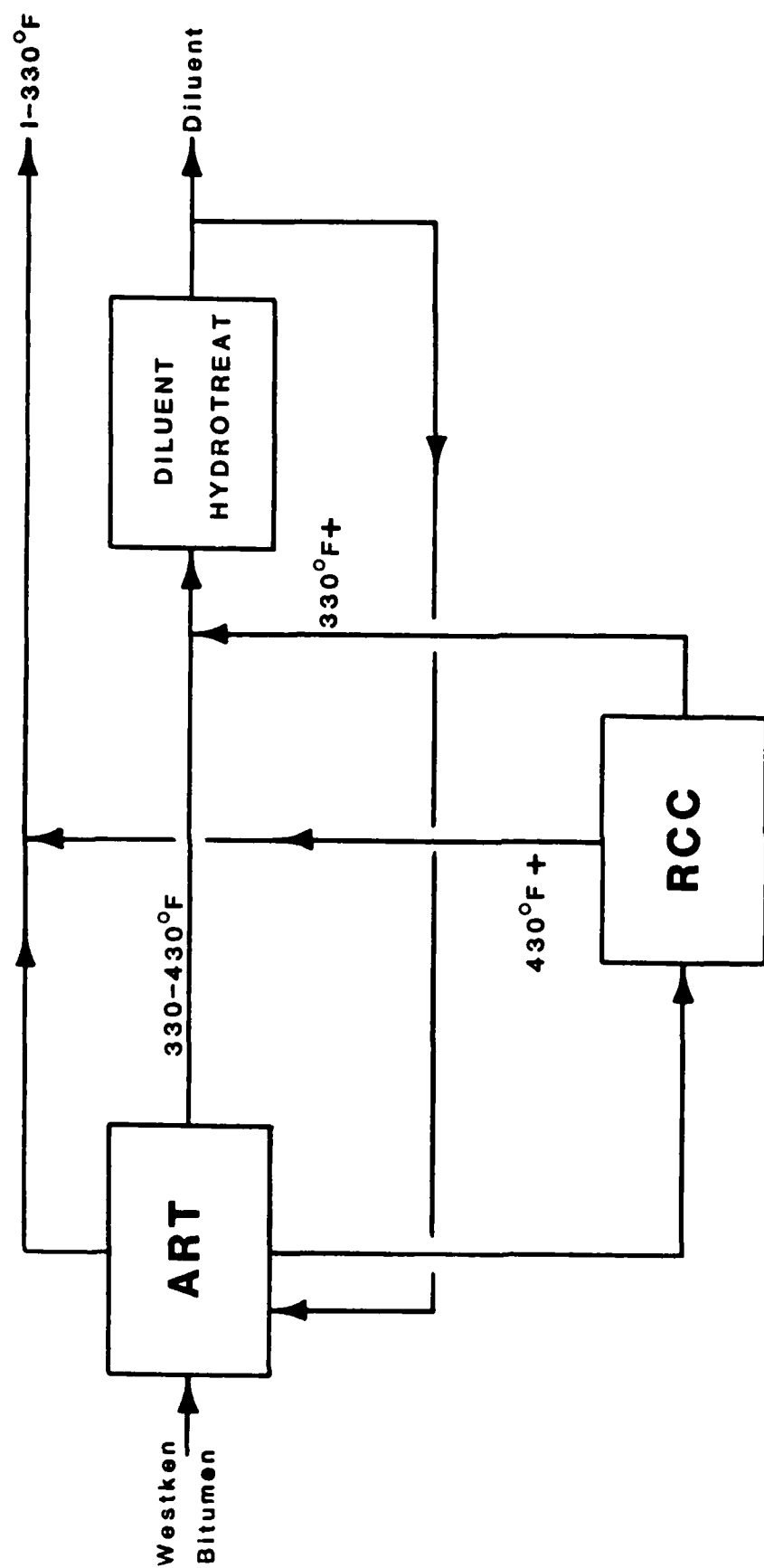


FIGURE 2. SIMPLIFIED SAMPLE PREPARATION CONVERSION SECTION FLOW DIAGRAM

TABLE 4

LOOP 1 BLENDED FEED TO THE ARTSM UNIT

Feed Blend Identification: 50/50 Blend by Weight of Westken
Bitumen and LCO Derived From
Westken Bitumen

Date of Blend: 4-15-85

Characterization

°API 17.7

ELEMENTAL ANALYSIS, WT%

HYDROGEN	10.74
SULFUR	0.87
TOTAL NITROGEN	0.100
BASIC NITROGEN	0.093
OXYGEN	1.73

VISCOSITY @ 210°F, CS 2.31

RAMSBOTTOM CARBON, WT% 3.75

POUR POINT, °F -10-

HPLC:

Saturates	33.9
Monoaromatics	22.1
Diaromatics	8.1
>Diaromatics	18.3
Polars	7.7
Asphaltenes	9.9

METALS:

Nickel, ppm	30
Vanadium, ppm	76
Iron, ppm	526
Sodium, ppm	426

TABLE 5

SUMMARY OF LOOP 1 ARTSM PROCESSING RESULTS

	<u>Week 1</u>	<u>Week 2</u>	<u>Diluent Only</u>
<u>Conditions</u>			
Sorbent:Oil Ratio	15.3	12.8	14.7
Temperature, °F	902	902	899
Water Injected, % Feed	17.6	13.0	11.0
<u>Yields, Wt% of Feed</u>			
Dry Gas	3.4	2.7	1.6
Wet Gas	3.0	2.0	1.6
C ₅ -430°F	17.4	17.2	23.8
430°F+	64.0	66.9	69.4
Coke	11.9	11.4	3.7

TABLE 6

SUMMARY OF LOOP 1 RCCSM RESULTS

	<u>Composite</u>	<u>Diluent Only</u>
<u>Conditions</u>		
Catalyst:Oil Ratio	12.2	13.9
Temperature, °F	968	971
Water Injected, Wt%	5.1	5.6
<u>Yields, Wt%</u>		
Dry Gas	2.4	1.6
Wet Gas	9.6	7.2
C ₅ -430°F	35.8	38.0
430°F+	46.6	48.8
Coke	5.6	4.4

transportation fuel yield was almost 65 wt% and occurred in a broad conversion range of 45 to 55 wt% conversion. The total cracked liquid product was composited and distilled into IBP-330°F and +330°F fractions. The 330°F+ material was blended with the Loop 1 330-430°F ARTSM product and hydro-treated in the pilot plant (2") hydrotreater over nickel-molybdate catalyst. Performance results are shown in Table 7. The performance of the catalyst is less than experienced during the diluent preparation experiments, probably due to the Loop 1 material being poorer quality (higher aromaticity).

LOOP 2 Conversion

Loop 2 followed a processing pattern similar to Loop 1. A mixture of Westken bitumen and hydrotreated diluent was desalted prior to ARTSM processing. The purpose of the desalting was to remove salt (particularly sodium) from bitumen prior to the ARTSM unit. This should be much more economical than depositing these metals on the ARTCAT sorbent. The Westken bitumen and diluent were fed to the Art process in a 1.6:1 weight ratio of bitumen to diluent. Sixty parts per million of Tretolite Tolad T-284 demulsifier was added and pH of the feed water adjusted to a pH of 8.

Salt removal ranged from 20 to 45%, well below what was expected. There was also poor separation of the water from the bitumen. The desalter product contained about 10%

TABLE 7

LOOP 1 DILUENT HYDROTREATING RESULTS

	<u>Feed</u>	<u>Product</u>
Gravity, °API	21.2	24.8
Elemental Analysis, Wt%		
Hydrogen	10.4	11.2
Sulfur	0.48	0.048
Nitrogen	0.059	0.011
Molecular Type, Wt%		
Saturates	-	36.1
Monoaromatics	-	45.7
Diaromatics	-	7.3
>Diaromatics	-	14.3
Polars		1.1
Distillation, °F at		
20%	382	387
50%	484	479
80%	664	635

water. These results clearly demonstrate that poor contacting and/or emulsion problems had occurred.

ARTSM treatment followed the desalting operation. The feed for this loop, (Table 8) was heavier than in Loop 1, to reduce the quantity of diluent and to improve process economics. Results show a higher than desired (42 percent) conversion due to high catalyst ratios, riser temperatures and water in the riser (due to excess water in the feed).

The ARTSM product was dewatered and distilled at 330°F and 430°F. The 430°F+ material was fed to the RCCSM cracking step, and the 330-430°F material was retained to blend with the 330°F+ RCCSM product for hydrotreatment. Two drums of 430°F+ ARTSM product were used for cracking (Table 9). Operations were comparable to Loop 1 except that the feed was poorer in quality due to the higher initial quantity of bitumen. The liquid product exclusive of cold trap material was distilled to produce an IBP-330°F cut and a 330°F+ cut with the 330°F+ material blended into the 330-430°F ARTSM product to provide feed for hydrotreating.

The diluent hydrotreater results (Table 10) were initially disappointing. The API gravity was increased from 19.8 to an average of 24.0 with a hydrogen consumption of 870 SCFB. A product containing 11.6 to 12.0% hydrogen content had been targeted, however, the hydrogen content of the product was 11.1 wt%. The poorer than anticipated results were due to the poorer quality feedstock produced from higher quantities

TABLE 8
SUMMARY OF LOOP 2 ARTSM OPERATION

<u>Feed Properties</u>		<u>Process Results</u>	
Gravity, °API	14.7	Conditions:	
Elemental Analysis, Wt%		Catalyst:Oil Ratio	22.8
Hydrogen	10.2	Temperature, °F	942
Sulfur	0.94	Water Injected, %	16.9
Nitrogen	0.092		
Ramsbottom Carbon	3.5	Yields, Wt%:	
Viscosity @ 210°F, cs	13.7	Dry Gas	2.9
Molecular Types, Wt%		Wet Gas	2.4
Saturates	24.7	C ₅ -430°F	21.4
Monoaromatics	23.0	430°F+	57.7
Diaromatics	15.0	Coke	15.6
>Diaromatics	16.7		
Polars	9.1		
Asphaltenes	11.5		

TABLE 9
LOOP 2 RCCSM OPERATIONS SUMMARY

<u>Feedstock</u>		<u>Process Results</u>	
Gravity, °API	17.8	Conditions:	
Elemental Analysis, Wt%		Catalyst:Oil Ratio	12.5
Hydrogen	11.0	Temperature, °F	971
Sulfur	0.6	Water Injected, %	5.0
Nitrogen	0.09		
Ramsbottom Carbon	2.5	Yield, Wt%:	
Viscosity @ 210°F, cs	3.01	Dry Gas	2.4
Molecular Types, Wt%		Wet Gas	9.1
Saturates	34.9	C ₅ -430°F	32.4
Monoaromatics	28.8	430°F+	49.9
Diaromatics	11.2	Coke	6.2
>Diaromatics	17.6		
Polars	7.1		
Asphaltenes	0.4		

TABLE 10

LOOP 2 HYDROTREATER RESULTS SUMMARYOPERATING CONDITIONS

Temperature, °F	697
Pressure, PSIG	1400
LHSV, Hr ⁻¹	1.4
Hydrogen Circulation, SCFB	3320

RESULTS

Liquid Yield, Wt%	100.0
Hydrogen Consumption, SCFB	870

LIQUID PROPERTIES

	<u>Feedstock</u>	<u>Product</u>
Gravity, °API	19.8	24.0
Hydrogen, Wt%	10.0	11.1
Sulfur, Wt%	0.45 (est)	0.02
Total Nitrogen, ppm	479	58
Basic Nitrogen, ppm	64	5
Sim-D, °F IBP	96	272
5%	325	324
10%	341	342
30	404	396
50%	479	451
70%	575	534
90%	743	670
95%	837	752
EP	1060	952

of bitumen in the initial loop feedstock. This hydrotreated cycle oil was then used as a jet fuel precursor in the final sample preparation.

Final Sample Preparation

The final fuel samples were prepared by olefin saturation of the naphtha, fractionating and hydrotreatment (Figure 3).

Feedstock blends were prepared for final JP-4 and JP-8 sample treating based on laboratory studies to determine the ratios needed to obtain appropriate jet fuel precursors. The objective of this work was to produce JP-4 and JP-8 hydrotreater feedstocks such that the precursors were consistent with the flow scheme and material balance, and all blends were representative of expected commercial unit intermediate products.

Based upon results from laboratory hydrotreating and Loop 2 diluent hydrotreating and fractionating, feeds were determined to be blends of the following:

JP-4: Olefin saturated naphtha and hydrotreated diluent fraction IBP-540°F.

JP-8: IBP-640°F hydrotreated diluent fraction.

The first step performed was diolefin saturation of the naphtha (I-330) blend components. This step was required because of the coking tendency and highly exothermic reaction

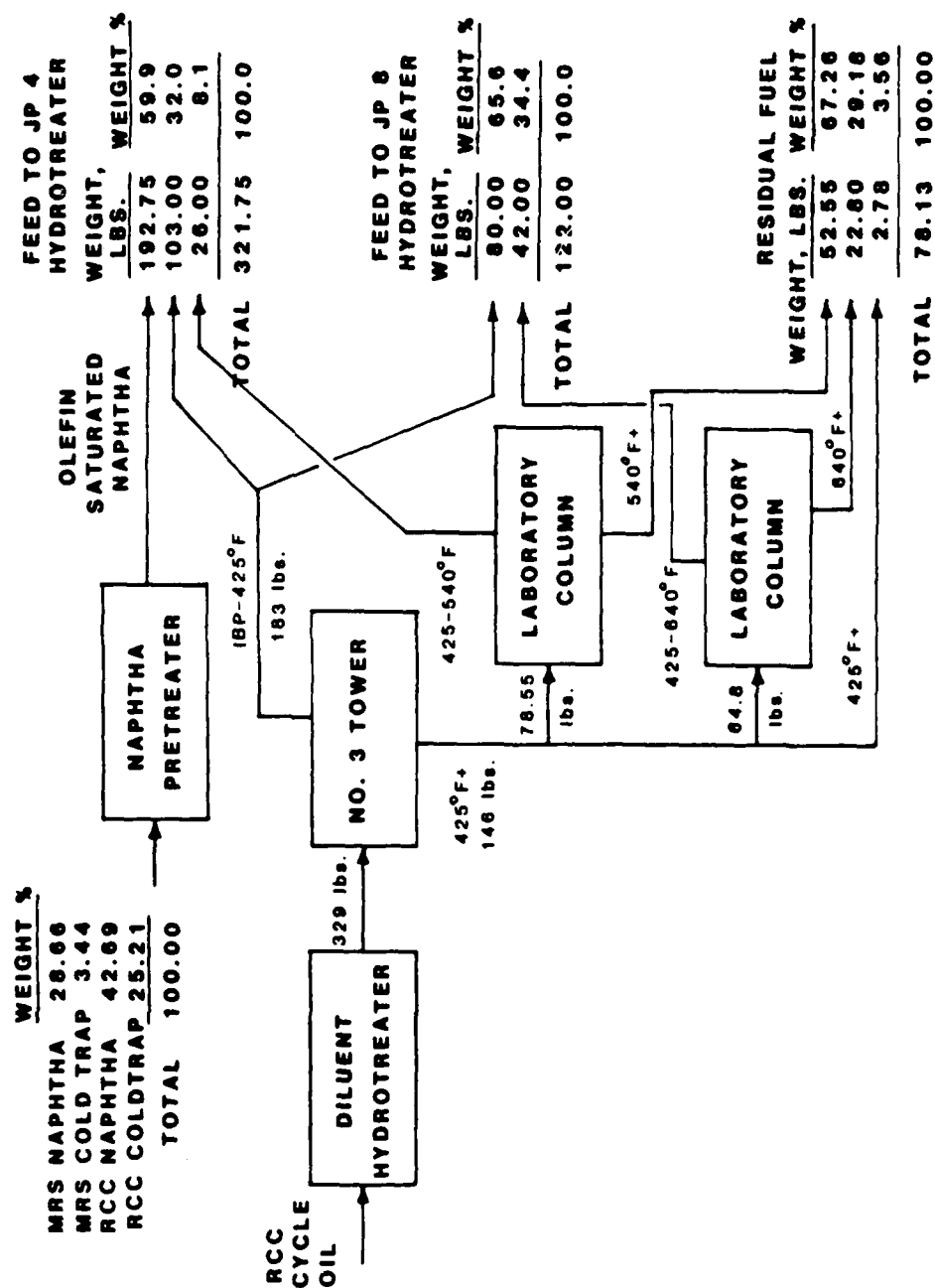


Figure 3. Turbine Fuel Precursor Blends

associated with saturation of diolefin components. This step required low severity, liquid phase hydrogenation. A two-pass (two-stage) operation was required to control the reactor exotherm (Table 11). Good performance was noted, but loss of light ends was encountered.

Distillation of the hydrotreated diluent was performed to produce the representative fractions, and product blends made as shown in Figure 3.

The blended JP-8 precursor was hydrotreated to produce specification JP-8 jet fuel using commercial nickel-molybdate catalyst. Overall results are shown in Table 12. The final sample met the gravity (40.0° API), hydrogen (13.57 wt.%), aromatics (12.0 vol%), sulfur (17 ppm) and distillation specifications. This sample did not meet corrosion specification (3b-4a) and required redistillation, caustic treating, and clay treating to reduce corrosiveness.

The final hydrotreating of the JP-4 fuels was performed with a commercial nickel-molybdate hydrotreating catalyst. After presulfiding, the reactors went through a 24 hour break-in procedure using cycle oil feed, which was then discarded. JP-4 hydrotreating conditions were intentionally varied to achieve variable levels of aromatics from 15 to 35% in the product. Table 13 summarizes these results. These fuel samples also did not meet corrosion specifications due to recombinant sulfur and required treatment by redistillation and clay treating to reduce the corrosion to acceptable levels.

TABLE 11

DIOLEFIN SATURATION
HYDROTREATMENT OF THE LIGHT NAPHTHA

<u>CONDITIONS</u>	<u>Feedstock</u>	<u>First Pass</u>	<u>Second Pass</u>
Temperature, °F		350	400
Pressure, PSIG		1400	1400
LHSV, Hr ⁻¹		7.6	3.6
Hydrogen Circulation, SCFB		1200	2300

LIQUID PROPERTIES

Gravity, °API	56.5	55.0	51.5
H ₂ , Wt%	12.57	12.85	14.07
Sulfur, ppm	450	-	250
Total Nitrogen, ppm	23	-	13

FIA, Vol%

Saturates	23.2	27.5	41.9
Olefins	53.9	41.9	25.3
Aromatics	22.9	30.6	32.8

Sim-D, °F @ Wt%

IBP/5	-14/68	-8/97	27/133
10/20	102/152	111/165	158/194
30/40	180/211	192/216	215/234
50	234	236	240
60/70	240/270	242/272	262/282
80/90	287/318	287/317	295/322
95/EP	329/356	324/394	333/482

TABLE 12

JP-8 HYDROTREATING SUMMARYPROCESS CONDITIONS

Temperature, °F	690
Pressure, PSIG	2000
LHSV, Hr ⁻¹	0.5
Hydrogen Circulation, SCFB	3900
Hydrogen Consumption, SCFB	1650
Liquid Yield, Wt%	101.4

LIQUID PROPERTIES

	<u>Feedstock</u>	<u>Product Average*</u>
Gravity, °API	27.6	40.0
Hydrogen, Wt%	11.25	13.6
Sulfur, ppm	176	17
Total Nitrogen, ppm	7	<1
Basic Nitrogen, ppm	1	<1
FIA - Vol%		
Saturates	22.6	87.1
Olefins	2.2	0.9
Aromatics	75.2	12.0
Sim-D, °F:		
20%	365	317
50%	425	383
80%	523	453

*Average analysis of six batch strippings

TABLE 13

JP-4 HYDROTREATING RESULTS SUMMARY

Aromatics Objective, %	15	25	30	35	
<u>Operating Conditions</u>					
Reactor Temperature, °F	685	609	617	564	
Reactor Pressure, PSIG	1200	1200	1200	1200	
LHSV, Hr ⁻¹	0.60	0.90	1.95	1.97	
Hydrogen Circulation, SCFB	3314	4380	3239	3209	
C ₅ ⁺ Liquids, Wt%	101.4	100.9	100.9	100.4	
Hydrogen Consumption, SCFB	968	742	557	407	
<u>Feedstock</u>					
Gravity, °API	43.0	48.2	46.8	45.1	44.1
Specific Gravity, gm/cc	0.8109	0.7874	0.7936	0.8014	0.8052
Hydrogen, Wt%	12.18	13.72	13.36	13.08	12.83
Sulfur, Wt% (ppm)	0.0515	(72)	(44)	(57)	(50)
N _T /N _B , ppm	10/4	<1	<1	<1	<1
FIA-Saturates	37.0	83.1	75.8	69.2	63.3
Olefins	1.7	0.7	0.7	0.8	0.8
Aromatics	61.3	16.2	23.5	30.0	35.9
Sim-D, °F @:					
20%	222	215	218	215	220
50%	299	291	305	298	304
80%	404	388	410	396	407
Copper Corrosion	-	-	4A	-	4B
Freeze Point	-	-90-	-	-	-

SECTION III

FINAL FUEL SAMPLE CHARACTERIZATION

Eight samples of military fuels were submitted from this program: one JP-8, four JP-4 samples, two gasolines, and one residual fuel.

One JP-8 aviation turbine fuel was submitted conforming to MIL-T-83133A specifications, (Table 14). Due to the characteristics of the Westken Tar Sands feedstock and the mode of processing utilized, the finished product was found to be highly naphthenic. The low API gravity, smoke point, low freezing point, and hydrogen content were indicative of the naphthenic character in contrast to a typical petroleum JP-8. Hydrogen content limitations required a 70°F reduction in the distillation end point of the fuel to meet specification. The smoke point was marginal due to the low hydrogen content. This fuel shows the high volumetric heating values of experimental "high density" fuels.

A total of four JP-4 aviation turbine fuel samples were submitted with sample 08-ND-133 conforming to MIL-T-5624L specifications, (Table 15). Three variable quality JP-4 samples having 25, 30 and 35 volume percent aromatics were also prepared for evaluation of the effects of higher aromatics contents on combustion. The present specification sample

TABLE 14.
FINAL TURBINE FUELS
JP8 MIL-T-83133A

<u>METHOD</u>	<u>SAMPLE NO. TEST</u>	<u>MIL- SPEC LIMIT</u>	<u>08-ND-132 Sample 1</u>
D156	COLOR (SAYBOLT)	Report	+30
D3242	ACIDITY, TOTAL(mg KOH/g)	0.015 Max	0.001
D1319	AROMATICS (VOL %)	25.0 Max	12.0
D1319	OLEFINS (VOL %)	5.0 Max	0.8
D1266	SULFUR, TOTAL (WT %)	0.3 Max	0.005
D86	DISTILLATION, INITIAL (°F)	Report	296
D86	10% REC.(°F)	401 Max	338
D86	20% REC.(°F)	Report	352
D86	50% REC.(°F)	Report	384
D86	90% REC.(°F)	Report	451
D86	FINAL BP(°F)	572	500
D86	RESIDUE (%)	1.5 Max	1.1
D86	LOSS (%)	1.5 Max	0.9
D93	FIASH POINT (°F)	100 Min	103
D1298	GRAVITY, API (60°F)	37-51	39.2
D1298	DENSITY, (Kg/l @ 15°C)	0.775-0.840	0.8289
D2386	FREEZING POINT (°F)	-58 Max	<-90
D445	VISCOSITY @ -4°F (cst)	8.0 Max	4.13
D3338	NET HEAT OF COMBUSTION, (Btu/Lb)	18,400 Min	18,505
D3343	HYDROGEN CONTENT (WT%)	13.5 Min	13.52
D1322	SMOKE POINT, mm	19 Min	19
	NAPHTHALENES, (VOL%)	3.0 Max	0.189
D130	COPPER STRIP (2 HR @ 212°F)	1B Max	1B
D3241	THERMAL STABILITY AT 500°F:		
	ΔP, mm Hg	25 Max	0.0
	PREHEATER TUBE COLOR CODE	2 Max	1
D381	EXISTENT GUM (mg/100 ml)	7 Max	1.8
D1094	WATER REACTION RATINGS	1B Max	1A
D3948	MSEP MODE A	*	93

*The minimum water separation index, modified, rating for JP8 shall be 85 with all additives except the corrosion inhibitor and the electrical conductivity additive, or 70 with all additives except the electrical conductivity additive.

TABLE 15.

FINAL TURBINE FUELS

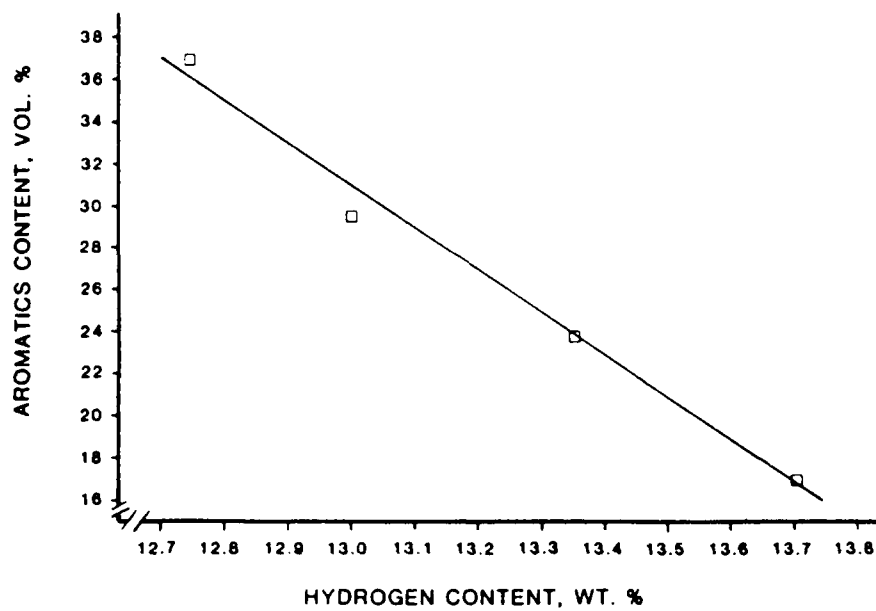
JP4 MIL-T-5624L

METHOD	SAMPLE NO. TARGET AROMATICS, VOLUME %	MIL- SPEC LIMIT	08-ND-133				08-ND-134				08-ND-135				08-ND-136			
			SAMPLE 1				SAMPLE 2				SAMPLE 3				SAMPLE 4			
			15				25				30				35			
			+30				+30				+30				+28			
			BC				BC				BC				BC			
D156	COLOR, SAYBOLT	Report																
D156	VISUAL BC-BRIGHT	Report																
	CL-CLOUDY																	
D3242	ACIDITY, TOTAL Mg KOH/g	0.015 Max	0.001				0.001				0.001				0.003			
D1319	AROMATICS, VOL %	25.0 Max	16.9				23.8				29.4				36.9			
D1319	OLEFINS, VOL %	5.0 Max	0.8				0.9				0.4				1.0			
D484	DOCTOR TEST (P-POSITIVE N-NEGATIVE)	Negative	N				N				N				N			
D1266	SULFUR, TOTAL (WT %)	0.40 Max	0.01				0.01				0.01				0.01			
D86	DISTILLATION, INITIAL (°F)	Report	156				160				163				156			
D86	10% REC. (°F)	Report	210				210				209				208			
D86	20% REC. (°F)	Max	232				232				224				234			
D86	50% REC. (°F)	Max	288				286				296				302			
D86	90% REC. (°F)	Max	405				403				427				448			
D86	FINAL BP (°F)	Max	478				486				500				512			
D86	RESIDUE (%)	1.5 Max	1.0				0.9				0.9				1.0			
D86	LOSS (%)	1.5 Max	1.0				1.1				1.1				1.0			
D1298	GRAVITY, API (60°)	45.0-57.0	47.6				46.8				44.5				43.0			
D1298	GRAVITY, SPECIFIC (60/60°F)	.802-.751	0.7901				0.7936				0.8040				0.8109			
D323	VAPOR PRESSURE, PSI	2.0-3.0	1.8				1.7				1.7				1.6			
D2386	FREEZING POINT (°F)	-72 Max	<-90				<-90				<-90				<-90			
D1405	ANILINE-GRAVITY PRODUCT	5250 Min	5259				4727				3960				3397			
D1322	SMOKE POINT, mm	20.0 Min	21.0				18.0				15.0				15.0			
D3343	HYDROGEN CONTENT (WT%)	13.6 Min	13.71				13.35				13.02				12.74			
D130	COPPER STRIP (2hr @ 212°F)	1B Max	1B				1B				1B				1B			
D3241	THERMAL STABILITY @ 500°F: ΔP, mm Hg	25 Max	0.0				0.0				0.0				0.0			
	PREHEATER TUBE COLOR CODE	<3	1				1				1				1			
D381	EXISTENT GUM (mg/100ml)	7.0 Max	1.0				1.2				4.8				5.8			
D1094	WATER REACTION RATINGS	1B Max	1A				1A				1A				1A			
D3948	MSEP-MODE B	70 Min	100				97				96				94			
	SAMPLE SIZE, GAL		9				10				9				10			

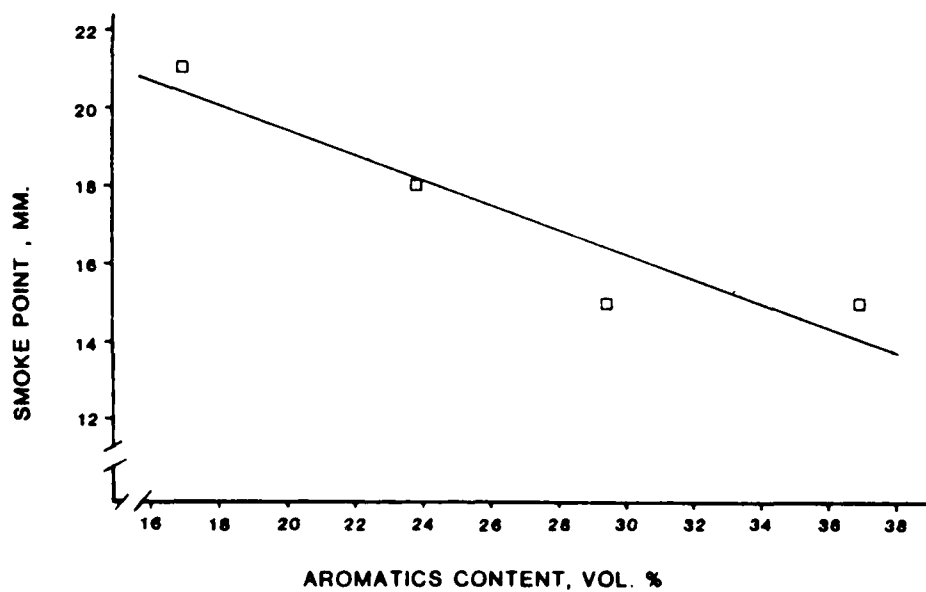
was an excellent fuel, meeting all required properties except volatility. The low vapor pressure, however, was due to sample handling rather than any process/fuel limitation. In contrast to conventional JP-4, this fuel was naphthenic, with a low API gravity, hydrogen content, and K factor. Thermal stability and freeze point were excellent. Key characteristics of all the JP-4 samples varied linearly with hydrogen and aromatics contents as shown in Figures 4 through 6.

The gasoline samples (Table 16) were high in olefins and aromatics with an excellent blend octane number. High copper strip corrosion values were due to elemental sulfur, remaining from the stripping of hydrogen sulfide.

The residual fuel oil (Table 17) represents an excellent, low sulfur content fuel. Due to its aromaticity, the gravity/viscosity relationship is somewhat different than for conventional residual fuels. Slight burner modifications or back-blending with raw bitumen would be required for direct use of this product.



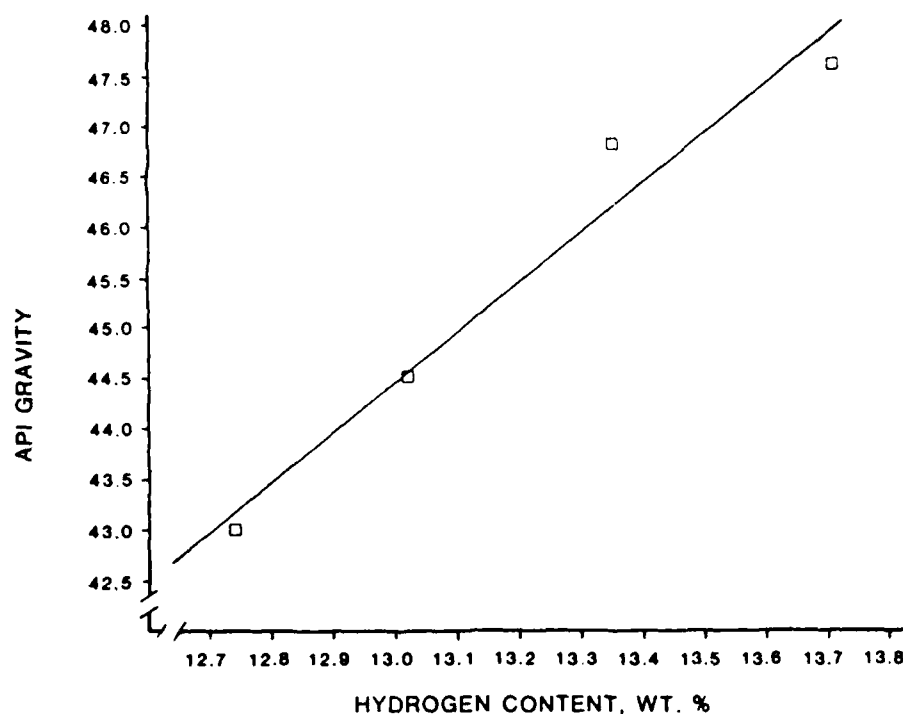
**Figure 4. Effect of Hydrotreating Severity
on JP-4 Aromatics Content**



**Figure 5. Effect of Hydrotreating Severity
on JP-4 Smoke Point**

TABLE 16.
GASOLINE SAMPLES

SAMPLE	LOOP 0 <u>08-ND-137</u>	LOOP 2 <u>08-ND-138</u>
Gravity, °API	56.2	61.4
Hydrogen, Wt. %	12.86	12.94
Sulfur, ppm	1400	603
Total Nitrogen, ppm	45	12
Bromine No.	127.5	115.5
RVP, psig	3.2	12.9
FIA, Vol. %		
Saturates	11.3	24.6
Olefins	64.4	40.9
Aromatics	24.3	34.5
Copper Corrosion	3B	4A
Octane No. (Blended)	107	109
Sim D: IBP °F	-8	-22
5%	87	31
10%	107	77
50%	234	209
90%	296	289
EP	338	335
Sample Size, Gallon	5	1



**Figure 6. Effect of Hydrotreating Severity
on JP-4 Density**

TABLE 17
FUEL OIL SAMPLE

	SAMPLE 08-ND-139	SPECIFICATIONS	
		#5 FUEL OIL	#6 FUEL OIL
Flash Point, F	230+	140 Min.	180 Min.
BS&W, Vol. %	Trace	1 Max.	1 Max.
Viscosity, SUS	113	125 Min.	900 Min.
@ 100 F		400 Max.	9000 Max.
Gravity, API	13.1	19 Typ.	13 Typ.
Sulfur, Wt%		Legal	Legal
Sample Size, Gal	1		

SECTION IV

DATA ANALYSIS

The primary objective of this task was to correlate all data developed in Phases I, II, and III, to determine commercial feasibility and projections, and to define any remaining problems and/or uncertainties associated with the upgrading processes. Data from each individual process module were compiled into usable data sets (desalting, ARTsm, RCCsm and Hydrotreating). Each data set was analyzed for correlation, accuracy, and fit with data from Phase I and Phase II.

Suitable variance of conditions was implemented in the ARTsm and RCCsm processing to obtain enough data for simple modeling without additional laboratory experimentation. Parameter variation runs were made at laboratory scale in order to firm up the hydrotreating response and predict conditions for the production runs of JP-4 and JP-8.

ARTsm and RCCsm data were processed to give yields based on 100% bitumen feed. Diluent contributions were mathematically backed out of the yields, thereby deleting the recycle effects. Smoothed data were then used to predict yields for typical commercial operating practice and these data were input to the computer optimization model.

Desalting

The Pilot Desalting Unit was qualified on crude oil prior to any treating of bitumen and essentially duplicated refinery operation on the same crude. Data from desalting the Loop 2 feedstock were used for evaluation of the desalting module. Laboratory data (Figure 7) at the same conditions of those of Loop 2 desalting were favorable; however, the pilot unit did not perform well even with demulsifier added. Subsequent runs in the pilot unit at different diluent dosages gave data as shown in Figure 8. Salt removal increased to a satisfactory rate; however, the large amount and type of diluent required to effect this rate would have a detrimental economic impact on the process. Since these data show desalting is possible, future work should include electrostatic precipitation as a possible means of oil/water separation at lower diluent dosages. Otherwise, desalting would have to be accomplished in the ARTSM unit at the price of higher adsorbent use.

For the purposes of this analysis, use of a desalting module was not practical since successful (commercially scaleable) desalting was not demonstrated. Based on prior experience, however, we would predict potentially successful desalting in a modern, multi-stage electrostatic unit with relatively high temperature and moderate dilution required.

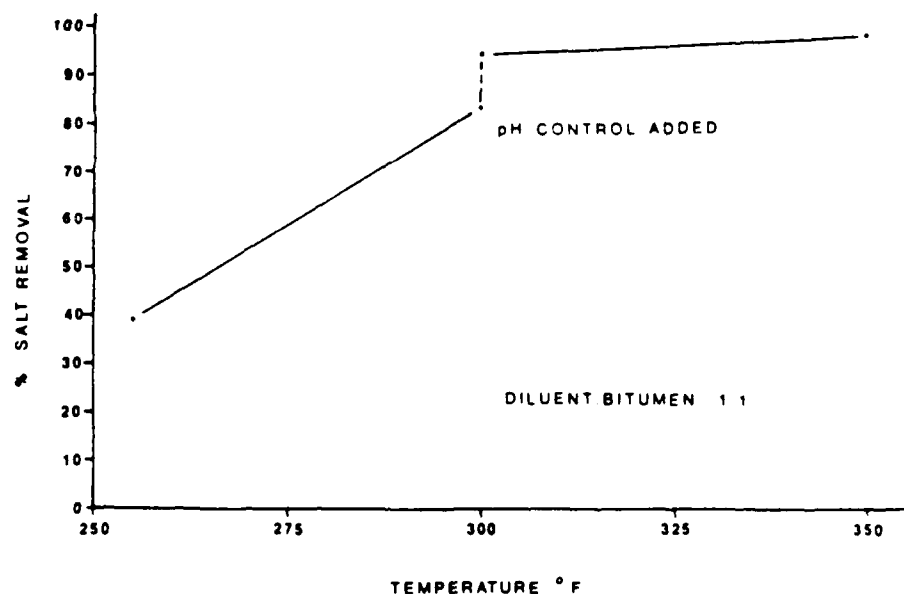


Figure 7. Laboratory Desalting Response

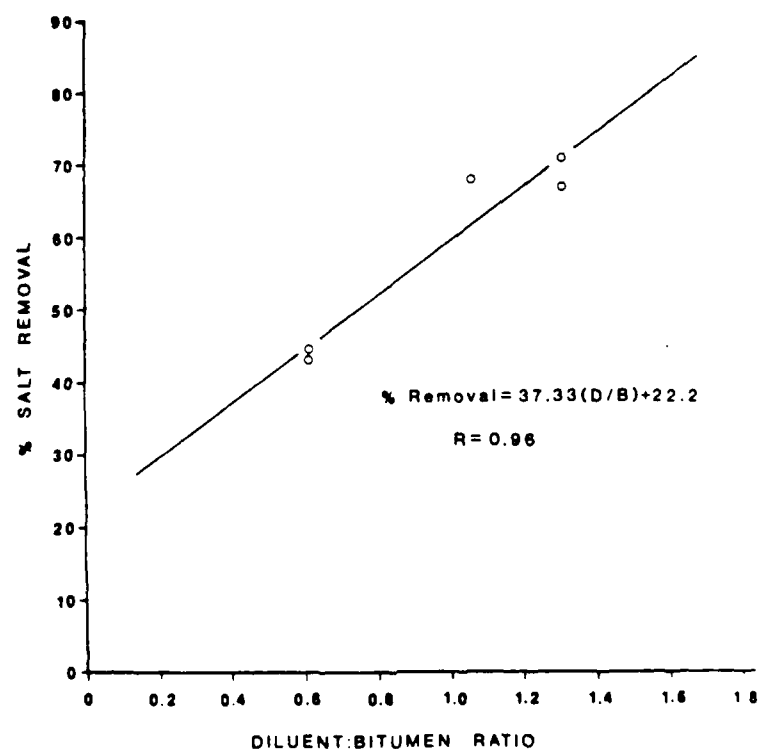


Figure 8. Pilot Scale Desalting Results

ARTSM

ARTSM processing proceeded satisfactorily, yielding less gas and naphtha with slightly higher distillate than had been predicted from Phase I and Phase II. This indicates less thermal cracking and is possibly due to the presence of the hydrogen donor recycle used as a diluent. Data from Loop 1 and Loop 2 processing were used to establish curves for product yields. The effects of the yields from the diluent alone were mathematically subtracted from each material balance so that only a bitumen response was left, shown in Figure 9.

In order to obtain the optimum commercial operating yields, the unit would normally operate at severities sufficient to produce a coke yield equivalent to the Ramsbottom carbon content of the feedstock. Current commercial operations at the Catlettsburg facility are within this region. Predicted yields for this feedstock are obtained from these curves, at a coke yield of 11%. The predicted yields are summarized in Table 18, compared to those predicted from Phase I and Phase II. Excellent agreement is shown with these earlier data, except for the decrease in gas yields.

RCCSM

RCCSM processing data were treated much in the same way as the ARTSM data. Data from Loop 1 and Loop 2 were used to establish

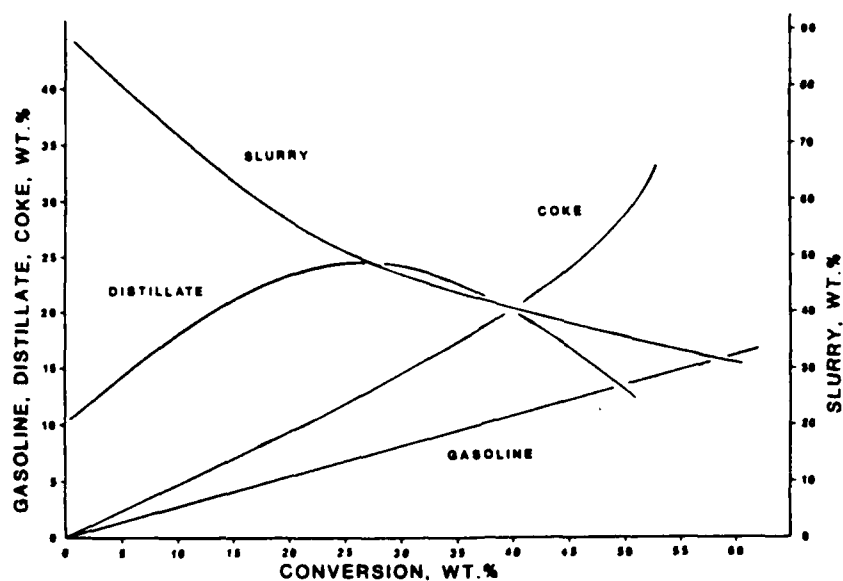


Figure 9. ART Module Yields for Diluent-Free Westken Bitumen

TABLE 18. PREDICTED COMMERCIAL ART MODULE YIELD STRUCTURE

COMPONENT	WEIGHT PERCENT OF FEED		
	DILUENT FREE BITUMEN BASIS		
	PHASE I	PHASE II	PHASE III
HYDROGEN	0.1	0.2	0.22
DRY GAS	3.3	2.4	1.96
C ₃ + C ₄	2.7	2.7	1.56
NAPHTHA	10.6	7.4	8.15
DISTILLATE	10.0/20.1*	23.2	23.80
SLURRY	62.7/52.6*	52.9	53.10
COKE	10.1	11.0	11.00
CONVERSION	27.3	23.9	23.10

* CORRECTED FOR DISTILLATE CONTENT OF BITUMEN

product yield response (Figure 10), again mathematically subtracting diluent yields. Figure 10 was used to determine the conversion level at which maximum total transportation fuels were produced. From this conversion level, the predicted yields for commercial operation were developed as shown in Table 19 and compared with predictions from Phase I and Phase II.

Phase III RCCSM processing yielded more naphtha and less slurry than had been previously predicted, with a higher conversion. In particular, more response in terms of catalytic yields (higher C₃+C₄ and gasoline, lower coke and slurry) were observed. These differences could be attributed to the presence of the hydrogen rich "donor" solvent recycle.

Diluent Hydrotreating

RCCSM cycle oil which was used as the diluent in Phase III was hydrotreated to partially saturate aromatics and to impart hydrogen donor properties to the stream. Loop 1 and Loop 2 cycle oil hydrotreating response data were used to develop a simple kinetic model. Results from the linearized model are shown in Figure 11. The relatively low temperature response shows the difficulty of hydrogenation of this material and low space velocities would be required to raise the hydrogen content markedly. The response to pressure is favorable and a good quality diluent can be produced at 1500 psig.

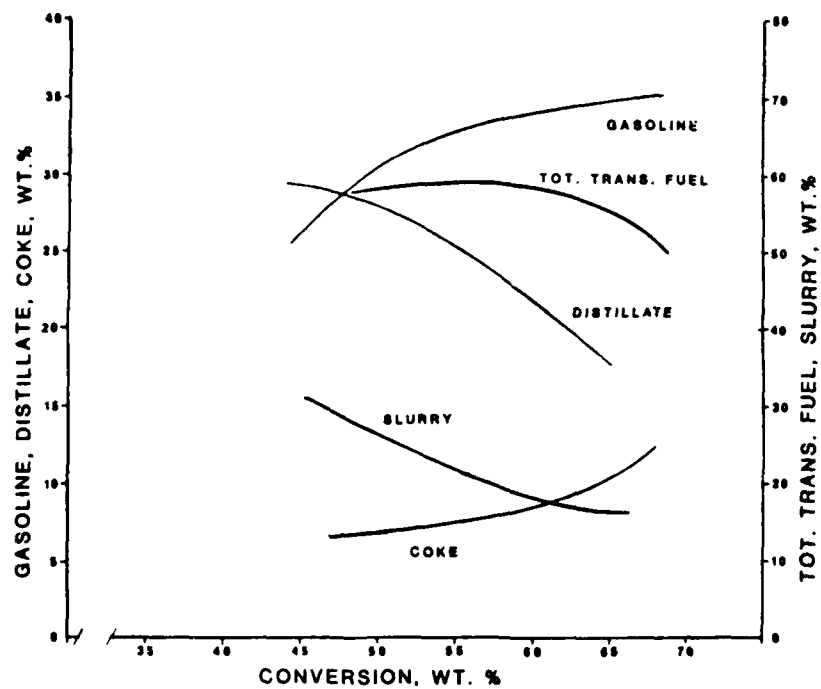


FIGURE 10. RCC YIELDS FOR DILUENT FREE WESTKEN BITUMEN

TABLE 19. PREDICTED COMMERCIAL RCC MODULE YIELD STRUCTURE

COMPONENT	WEIGHT PERCENT OF FEED		
	DILUENT FREE BITUMEN BASIS		
	PHASE I	PHASE II	PHASE III
HYDROGEN	0.10	0.33	0.18
DRY GAS	1.89	3.33	2.56
C ₃ + C ₄	9.95	10.38	11.33
NAPHTHA	30.10	15.10	33.51
DISTILLATE	18.42	37.22	23.51
SLURRY	34.32	19.82	20.51
COKE	5.01	13.42	7.20
CONVERSION	48.53	42.96	57.50

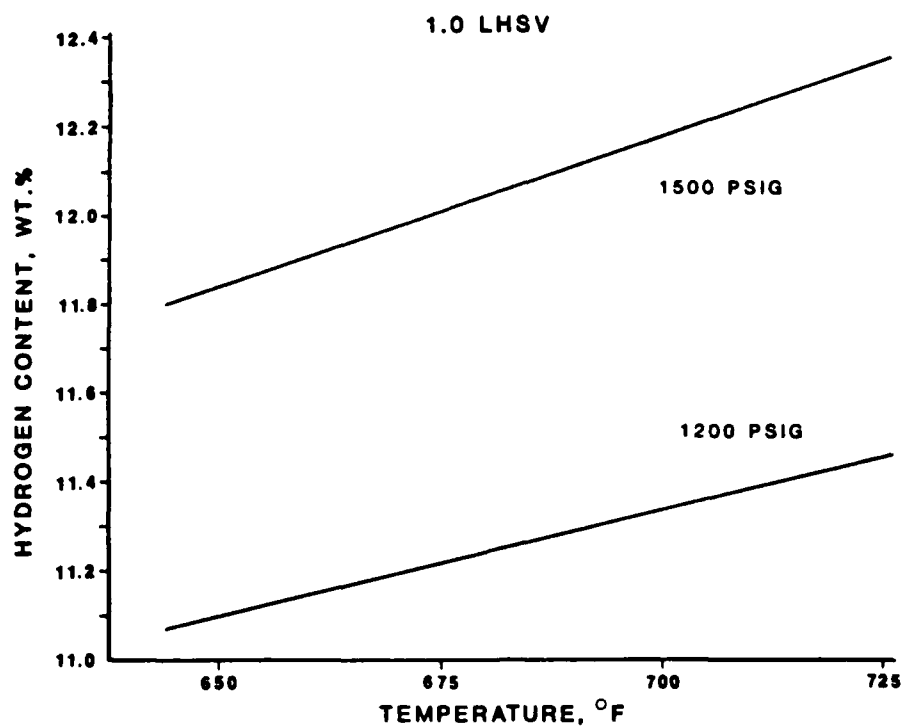


FIGURE 11. DILUENT RESPONSE TO HYDROTREATING CONDITIONS

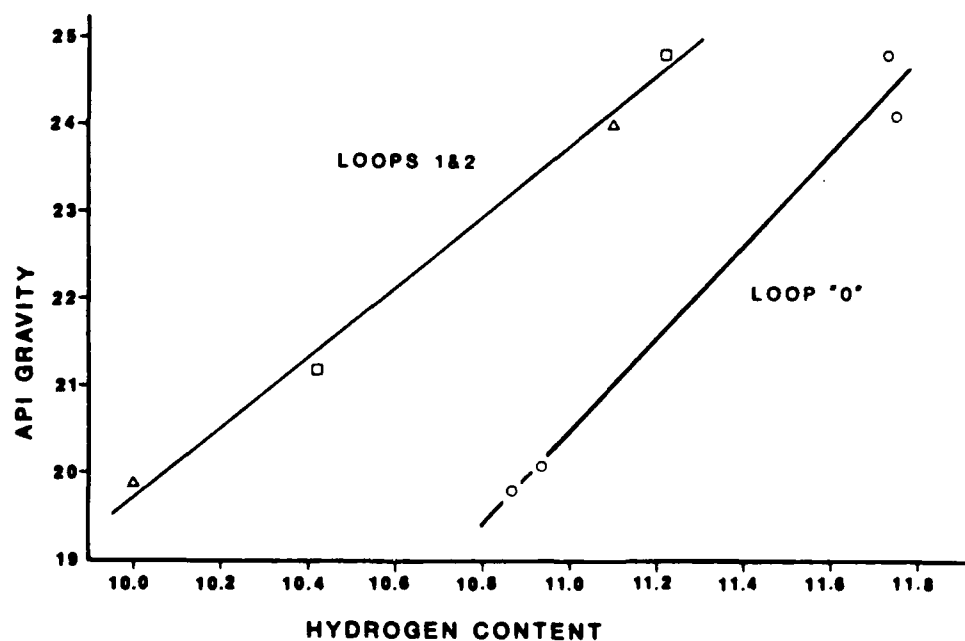


FIGURE 12. COMPARISON OF HYDROTREATED DILUENT PROPERTIES

As noted previously, multiple preparation "loops" were provided to allow approach to recycle convergence. A diluent was prepared in Loop "0" by processing Westken gas oil diluted bitumen through the RCCSM pilot unit, with Loop 1 and Loop 2 cycle oil derived directly from the process. Figure 12 shows the descriptive properties of the Loop "0" diluent compared with those of the diluent from Loops 1 and 2. The differences in hydrogen content and API gravities demonstrate the different nature of the diluents. Loops 1 and 2 recycle properties appear to have converged, as was hoped.

JP-4 Hydrotreating

JP-4 hydrotreating was predictable, routine and not as difficult as originally thought. A lower pressure was required in Phase III than in Phase II to obtain an on-specification product. Table 20 compares conditions and results from Phase II and Phase III processing. Feedstock differences between phases are indicated by gravity, boiling range and hydrogen content.

Hydrotreating response is shown by the linearized model of Figure 13. Processing at 1.0 LHSV was shown to be marginal for the 13.6 wt% specification hydrogen content. A 0.6 to 0.8 LHSV allowed reactor temperature to remain below 700°F and hydrogen partial pressure below 1200 psig.

TABLE 20. JP-4 HYDROTREATING DATA SUMMARY

CONDITIONS	PHASE II	PHASE III
TEMPERATURE, °F	685	685
PRESSURE, PSIG	2000	1200
LHSV, 1/HR.	0.5	0.6
HYDROGEN RATE, SCFB	5000	3300
PRODUCT PROPERTIES		
GRAVITY, °API	53.2	48.2
HYDROGEN, WT%	14.12	13.72
AROMATICS, VOL%	1.9	16.2
50% WT. TBP, °F	357	291
FEED PROPERTIES		
GRAVITY, °API	33.7	43.0
HYDROGEN, WT%	10.74	12.18
AROMATICS, VOL%	61.6	61.3
50% WT. TBP, °F	396	299

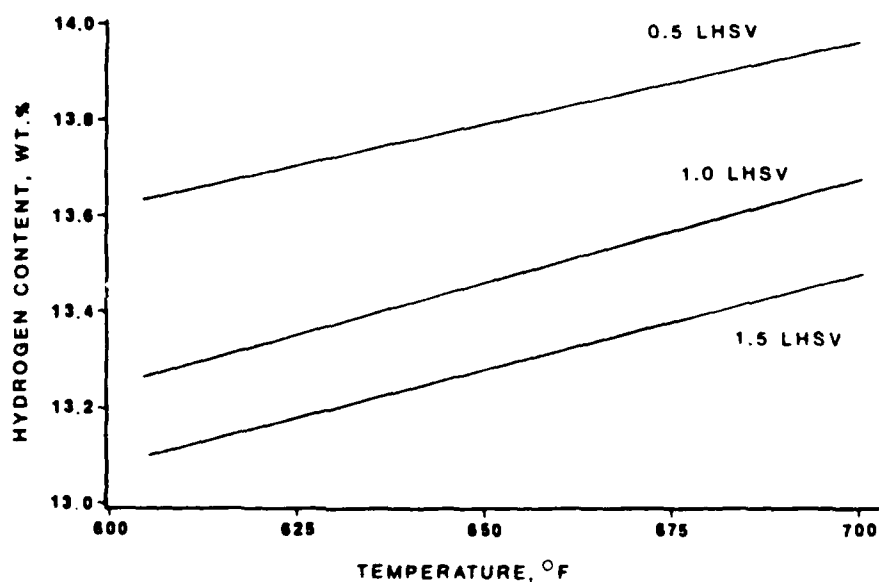


FIGURE 13. JP-4 HYDROTREATING RESPONSE AT 1200 PSIG

JP-8 Hydrotreating

JP-8 hydrotreated with much more difficulty than did JP-4. A preliminary laboratory parameter variation study showed that an extremely low LHSV (0.2) would be required to give the necessary 2.3 wt% increase in hydrogen content at 700°F and 1200 psig. Pressure effects gave a better response, however, and a satisfactory hydrogen content was obtained at 2000 psig, 700°F and 0.5 LHSV. Table 21 compares the results of the final JP-8 hydrotreating for Phase II and Phase III. As previously noted, the boiling range of the Phase III sample had been reduced in order to meet the hydrogen specification. Figure 14 depicts the difficulty of hydrotreatment of the Phase III sample. Although the aromatics specification is easily met, the final product is highly naphthenic and slightly on the hydrogen deficient side. It remains, however, a high quality turbine fuel.

Overall, the data analysis showed a relatively good comparison between Phase II and Phase III. Differences proved to be positive and explainable in view of the hydrotreated diluent used in Phase III.

TABLE 21. JP-8 HYDROTREATING DATA SUMMARY

	PHASE II	PHASE III
CONDITIONS		
TEMPERATURE, °F	700	690
PRESSURE, PSIG	2000	2000
LHSV, 1/HR.	0.5	0.5
HYDROGEN RATE, SCFB	5000	3800
PRODUCT PROPERTIES		
GRAVITY, °API	37.9	40.0
HYDROGEN, WT%	13.68	13.6
AROMATICS, VOL%	4.1	12.0
50% WT. TBP, °F	412	383
FEED PROPERTIES		
GRAVITY, °API	19.2	27.6
HYDROGEN, WT%	9.57	11.25
AROMATICS, VOL%	90.9	75.2
50% WT. TBP, °F	472 EST.	425

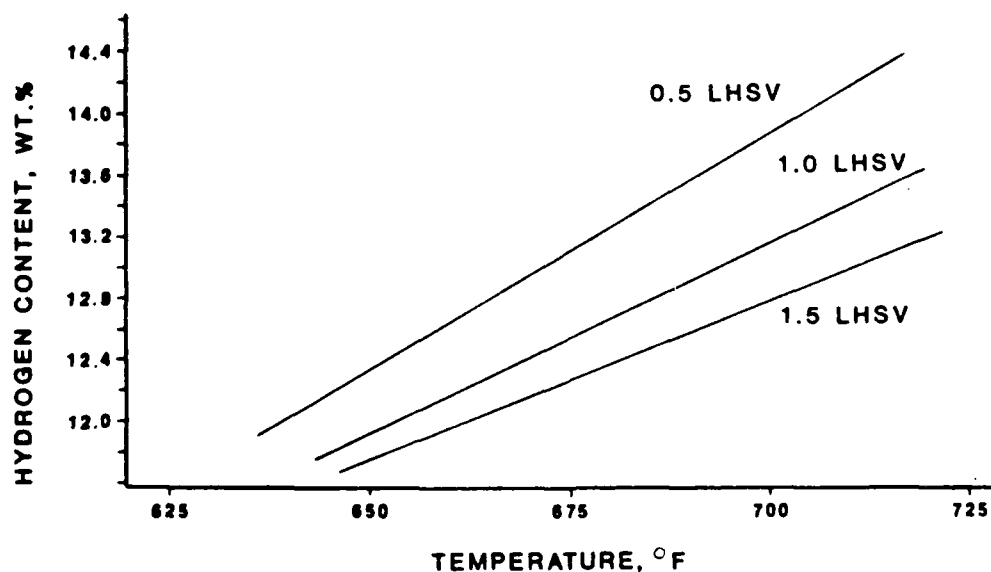


FIGURE 14. JP-8 HYDROTREATING RESPONSE AT 2000 PSIG

SECTION V

ECONOMIC RESULTS

Potential project economics for processing Westken bitumen in the mode successfully demonstrated in this program have been developed. An LP model based on these data has been utilized to develop project economics for several scenarios.

Bases and Assumptions

Assumptions and bases used in these studies were defined in conjunction with Air Force personnel, as detailed in Table 22. These values were selected to be representative at the date of the original study.

Capital costs were estimated by two methods. RCCsm/ARTsm capital costs were based on 1983 construction of a 55,000 BPD ARTsm unit and a 40,000 BPD RCCsm unit at Catlettsburg, Kentucky. Total base costs for this complex were approximately 300 million dollars, including process units, main columns, gas concentration, limestone boilers, baghouses and major supply systems. Capital costs for other plant sections were obtained from literature values. All values were updated to a Chemical Engineering cost index of 326, and off-site facilities were evaluated as 45% of plant on-sites.

TABLE 22. SUMMARY ECONOMIC BASES AND ASSUMPTIONS

CRUDE INVENTORY: 21 DAYS STORAGE CAPACITY/14 DAY INVENTORY.	
PRODUCT INVENTORY: 14 DAYS STORAGE CAPACITY/7 DAY INVENTORY.	
CRUDE MATERIAL: \$20/BBL	
PRODUCT PRICE: 0 ALL LIQUID MILITARY TRANSPORTATION FUELS, GASOLINE, JP-4, JP-5, JP-8, DF-2, VALUED AT EQUAL VALUE AS CALCULATED FOR A 15% DCF RATE OF RETURN.	PLANT LOCATION: MIDWEST
0 FUEL GAS \$20.00/FOE BBL	REFINERY CAPACITY: 25,000 BPD
0 PROPANE \$16.00/BBL	COST BASE: CE INDEX = 326
0 ISO BUTANE \$31.00/BBL	PLANT OFF-SITES: 45% OF PLANT ON-SITES EXCLUSIVE OF SPECIFIED FEED AND PRODUCT TANKAGE
0 NORMAL BUTANE \$29.00/BBL	FINANCING: - 100% EQUITY
0 AMMONIA, ANHYDROUS \$210.00/SHORT TON	- THREE-YEAR PLANT CONSTRUCTION PERIOD
0 SULFUR \$125.00/LONG TON	25% 1ST YEAR
0 RESIDUAL FUEL OIL \$20.00/ BBL	50% 2ND YEAR
	25% 3RD YEAR
DEBT FINANCING: 15%	INVESTMENT TAX CREDIT: 10% 1ST YEAR
PROCESS HEAT: \$20.00/BBL FOE	DISCOUNTED CASH FLOW RATE: 15%
COOLING WATER: \$.07/1000 GALLONS	PLANT SALVAGE VALUE: ZERO
BOILER FEED WATER: \$.40/1000 POUNDS	PLANT DEPRECIATION: 5 YEAR ACCELERATED COST RECOVERY SYSTEM.
ELECTRICAL POWER: \$.05/KWHR	PLANT LIFE: 16 YEARS
OPERATOR: \$16.00/MANHOURL	PLANT OPERATING FACTORS: 50% CAPACITY 1ST YEAR
HELPERS: \$14.00/MANHOURL	PLANT ON STREAM FACTOR: 90% AFTER 1ST YEAR
SUPERVISION: 25% OF DIRECT LABOR	STARTUP COSTS: 10% OF ESTIMATED ERECTED PLANT COSTS
OVERHEAD: 100% OF DIRECT LABOR	
TAXES: FEDERAL AND STATE COMBINED @ 50%	
MAINTENANCE, TAXES, INSURANCE: 4.5% OF FIXED INVESTMENT	

Plant sizes were selected to provide a total of 50,000 BPD of total feed to the demetallation section, limiting total bitumen input to 25,000 BPD due to the requirement of a diluent for proper feed distribution and fluidization. Plant sizes were originally selected to be near-optimum scale for single plant ARTSM modules. Larger scale units (possibly with multi-train ARTSM units) would decrease the plant capital costs per barrel of throughput and therefore reduce final product costs.

Operating costs and feedstock values were estimated at mid-1986 levels. Product value calculations were based on equal-value transportation fuels at a 15% DCF rate of return. Transportation fuels were defined as gasoline, diesel, JP-4, and JP-8. All other plant products were valued as byproducts.

Modeling and Case Studies

Data developed in Phase I, II, and III were used to develop an overall refinery LP model for these materials. The model provided for processing and blending materials to conventional specification fuels, such that all required constraints and product requirements were met. Only conventional finished materials were allowed, while inputs were limited to the Westken bitumen, isobutane, normal butane, and electrical power.

Figure 15 shows the major flow options allowed in the case study analyses. This flow scheme is based on actual results

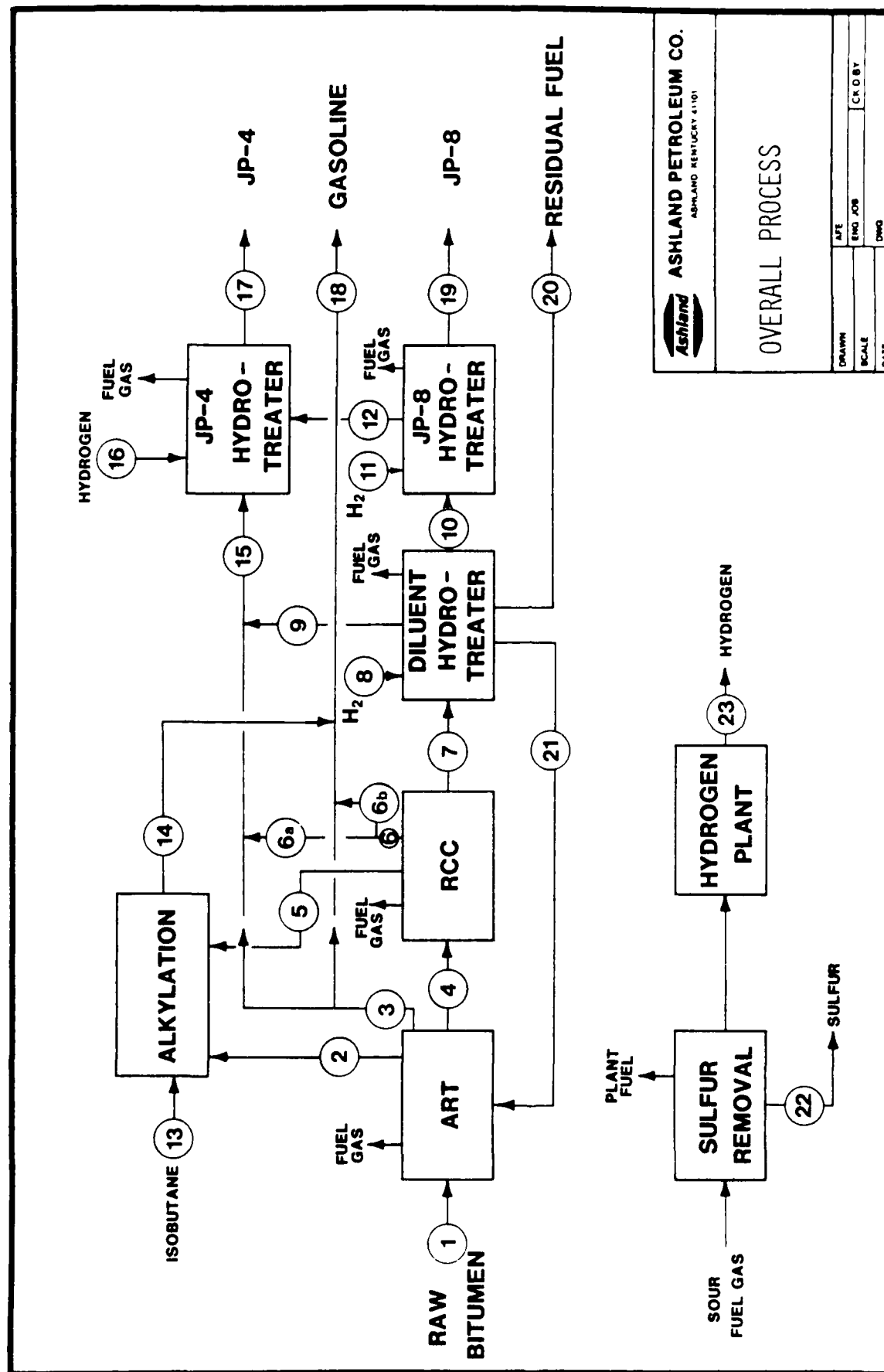


FIGURE 15. MAJOR PROCESS FLOW OPTIONS

obtained during this program, without allowance for potentially improved routes for which no data were available. Only process modules demonstrated during the experimental effort were used in the analysis, and only actual, measured process response data for ARTSM, RCCSM, and hydrotreater modules were utilized. The model was constrained to use 25,000 barrels per day of bitumen, but allowed to make any product slate with the overall goal of profit optimization. As a result, process modules and/or product slates are changeable from case to case. A base case was prepared using present specifications and requirements, with change cases used to define differential responses as listed in Table 23. Detailed flow sheets for major cases, and major flow quantities, are included in Appendix B.

Base Case

The base case, Table 24, was defined as an open product slate, profit-optimized plant producing only conventional fuels. Total fuel yield was 90 volume percent, or 86.9 volume percent transportation fuels. Net thermal efficiency was only 79+%, suggesting excess coke or fuel production within the plant boundaries.

TABLE 23. PHASE III CASE STUDIES

<u>CASE NUMBER</u>	<u>DESCRIPTION</u>
1	BASE CASE, OPEN COST OPTIMUM SOLUTION, ALL PRESENT SPECIFICATIONS.
1A	SAME AS BASE, EXCEPT 75% DEBT/25% EQUITY.
1C	SAME AS BASE, EXCEPT 7500 BPD BITUMEN FEED RATE.
2	EXTEND JP-8 SPECIFICATIONS TO 0.865 SPECIFIC GRAVITY AND 13.4% HYDROGEN.
3	EXTEND JP-8 SPECIFICATIONS TO 13.3% HYDROGEN.
4	EXTEND JP-8 SPECIFICATIONS TO 13.0% HYDROGEN.
5	MAXIMUM TURBINE FUEL - HIGH VALUE DIFFERENTIAL ALLOWED FOR BOTH JP-4 AND JP-8.
6	MAXIMUM JP-4--HIGH VALUE DIFFERENTIAL ALLOWED FOR JP-4.
7	MAXIMUM JP-8--HIGH VALUE DIFFERENTIAL ALLOWED FOR JP-8 WITH NORMAL SPECIFICATIONS.
8	MAXIMUM JP-8--HIGH VALUE DIFFERENTIAL ALLOWED FOR JP-8 WITH 0.865 SPECIFIC GRAVITY AND 13.0% HYDROGEN SPECIFICATION EXTENSIONS.

TABLE 24. SUMMARY OF BASE CASE ECONOMIC RESULTS

Case Number:	1			
Case Description:	Base Case, All Present Specification			
Feeds:	BPD	TPD	Vol%	Wt%
Bitumen	25000	4363	75.2%	84.2%
Isobutane	6202	611	18.7%	11.8%
Normal Butane	2035	208	6.1%	4.0%
Subtotal Feeds	33237	5182	100.0%	100.0%
Products, BPD:				
Propane	980	87	3.3%	2.2%
Unleaded	25117	3225	83.9%	83.2%
JP-4	0	0	0.0%	0.0%
JP-8	3782	554	12.6%	14.3%
Residual Fuel	43	8	0.1%	0.2%
Subtotal Liquids Out	29922	3874	100.0%	100.0%
Sulfur, TPD		51		
Yield, vol % of feeds	90.0%		Thermal Efficiency	
Vol. % Transportation fuel	86.9%		Net	79.5%

CAPITAL INVESTMENT:

Unit	Cost, MM\$	Percent of Total
ART	107.9	28.1%
RCC	88.9	23.2%
Recycle Hydrotreater	12.2	3.2%
Naphtha Pretreater	0.0	0.0%
JP-4 Hydrotreater	0.0	0.0%
JP-8 Hydrotreater	16.2	4.2%
Alkylation	14.5	3.8%
Hydrogen Plant	12.6	3.3%
Sulfur Plant	4.2	1.1%
Subtotal Battery Limits	256.6	66.9%
Tankage	11.8	3.1%
Offsites @ 45%	115.5	30.1%
Fixed Capital Investment	383.8	100.0%

TRANSPORTATION FUEL COST ELEMENTS:

ITEM	COST, \$/bbl	PERCENT OF TOTAL
Startup	0.23	0.5%
Working Capital	0.29	0.7%
Byproducts	-0.75	-1.7%
Fixed Costs	2.55	5.8%
Income Taxes	3.77	8.6%
Utilities	5.15	11.7%
Capital Related	6.59	15.0%
Raw Materials	26.00	59.3%
Prime Product Cost, \$/bbl	43.83	100.0%

Primary product from the plant was gasoline. Since profit optimization was used to define product slates, favored products were low hydrogen content, low "degree of processing" materials, e.g., cracked gasoline. In fact, a major advantage of this process was production of large quantities of low hydrogen content materials. Increasing the hydrogen content to turbine fuel requirements would increase costs proportionately.

Table 25 summarizes capital costs for the plant. The relatively large size of the ARTSM and RCCSM units resulted in these modules comprising the major portion of plant capital.

Startup, working capital, and byproduct credits (for LPG, sulfur, and residual fuel) were minor contributions to total cost. The major cost element was raw material, with capital, utilities, income taxes, and fixed costs representing lower elements by an order of magnitude.

In comparison, Table 26 summarizes Phases I, II, and III results. The Phase III results were significantly improved over Phase II, with this difference primarily attributed to the use of the hydrogen enriched recycle stream. The low hydrogen content of the Westken material obviously requires hydrogen input early in this process. In fact, hydrotreatment of the ARTSM product (prior to RCCSM) may well be favored over the present route. Conversely, Phase III results were slightly poorer than Phase I predictions. This again is probably due to the relatively refractory nature of the feed,

TABLE 25. SUMMARY ECONOMIC RESULTS COMPARISON

	PHASE I	PHASE II	PHASE III
INVESTMENT DATA, MM\$			
FIXED CAPITAL	360	436	384
WORKING CAPITAL	25	19	20
MATERIAL FLOWS, BPCD			
INPUTS:			
BITUMEN	29999	25000	25000
ISOBUTANE	4634	2094	6202
NORMAL BUTANE	2144	998	2035
PRODUCTS:			
LPG	1240	191	980
GASOLINE	25979	19038	25117
JP-4	2524	2123	0
DIESEL FUEL/JP-8	3461	3101	3782
RESIDUAL FUEL	630	-	43
PRODUCT COST \$/BBL AT 15% DCF:			
STARTUP	0.20	0.28	0.23
WORKING CAPITAL	0.34	0.27	0.29
BYPRODUCTS	(1.46)	(0.37)	(0.75)
FIXED COSTS	2.04	2.83	2.55
INCOME TAXES	3.22	5.15	3.77
UTILITIES	3.31	4.82	5.15
CAPITAL	5.62	8.89	6.59
RAW MATERIALS	29.90	29.63	26.00
PRIME PRODUCT COST, \$/BBL	43.20 =====	51.50 =====	43.83 =====

TABLE 26. THE EFFECT OF FINANCING METHOD AND PLANT SIZE ON THE
BASE CASE PLANT PARAMETERS.

CASE NUMBER	1	1A	1C
CASE NAME	Base	75% Debt	7500 BPD
PLANT FLOWS, BPCD:			
Feeds:			
Butumen	25000	25000	7500
Isobutane	6202	6202	1861
Normal butane	2035	2035	610
	-----	-----	-----
Subtotal Feeds:	33237	33237	9971
	-----	-----	-----
Products:			
Propane	980	980	294
Unleaded Gasoline	25117	25117	7535
JP-4	0	0	0
JP-8	3782	3782	1135
Residual Fuel	43	43	13
	-----	-----	-----
Subtotal Liquids	29922	29922	8977
	-----	-----	-----
Sulfur, TPCD	51	51	15
	=====	=====	=====
Yields, Volume %:			
Total Liquids	90.0%	90.0%	90.0%
Transportation Fuel	86.9%	86.9%	86.9%

CAPITAL INVESTMENT, MM\$:			
ART	107.9	107.9	52.4
RCC	88.9	88.9	43.2
Recycle Hydrotreater	12.2	12.2	5.7
Naphtha Pretreater	0	0	0
JP-4 Hydrotreater	0	0	0
JP-8 Hydrotreater	16.2	16.2	7.4
Alkylation	14.5	14.5	6.9
Hydrogen Plant	12.6	12.6	5.1
Sulfur Plant	4.2	4.2	2.1
	-----	-----	-----
Battery Limits	256.5	256.5	122.8
Tankage	11.8	11.8	5.5
Offsites at 45%	115.5	115.5	55.2
	=====	=====	=====
Fixed Investment	383.8	383.8	183.5
	=====	=====	=====

OPERATING COSTS, MM\$/yr:			
Utilities	54.3	54.3	16.3
Fixed Costs	26.1	26.1	17.0
Byproduct Credits	-7.9	-7.9	-2.4
	-----	-----	-----
Net Operating Costs	72.5	72.5	31.0
	-----	-----	-----

and the need for early hydrogen enrichment which was not originally anticipated.

Table 27 summarizes the impacts of plant size and accounting method on the base case plant costs. Using 75% debt financing with all other factors constant, transportation fuel costs were reduced by over \$3.00/barrel. The primary impact of this option was reduction of total income taxes paid.

TABLE 27. THE EFFECT OF FINANCING METHOD AND PLANT SIZE ON TRIAL PRODUCT COSTS

CASE NUMBER	1	1A	1C
CASE NAME	Base	75% Debt	7500 BPD
TRANSPORTATION FUEL COST COMPONENTS, \$/bbl:			
Startup	0.23	0.23	0.37
Working Capital	0.29	0.29	0.31
Byproducts	-0.75	-0.75	-0.75
Fixed Costs	2.55	2.55	5.56
Income Taxes	3.77	0.43	6.05
Utilities	5.15	5.15	5.15
Debt Service	0.00	5.26	0.00
Capital	6.59	1.47	10.53
Raw Materials	26.00	26.00	26.00
	-----	-----	-----
Total Cost, \$/bbl	43.83	40.62	53.23
	=====	=====	=====

Reducing plant size to 7500 BPD could be considered for a demonstration, site-specific project. An integrated plant of this size is definitely not economically attractive; on-site upgrading to a synfuel and sale to a remote refinery would be much more feasible. However, for discussion purposes, all

costs were scaled to 7500 BPD of bitumen. Because of the amount of scale reduction, the uncertainty in capital costs rises significantly and these values should be used with caution. The major cost impacts of this change were significant increases in the fixed and capital cost contributions.

Hydrogen Content of JP-8

The base case produced about 3800 BPD of conventional specification JP-8 fuel. Due to the naphthenic nature of this fuel, the boiling range of the fuel had to be reduced significantly to meet the 13.5% hydrogen specification. In order to evaluate the effects of this constraint, incremental reductions in the hydrogen content specification and a specific gravity increase were evaluated in terms of plant operation and product costs, (Table 28).

Overall, as the hydrogen specification was reduced, hydrogen content of the final fuel was lowered an equal amount. Of particular interest, the fuel became heavier as higher-boiling components previously restricted by hydrogen content displaced lighter components into the gasoline pool. Total plant production increased in this case due to lower severity processing requirements, and this lower severity operation was reflected in lower plant capital costs. Actual JP-8 production, however, decreased.

TABLE 28. THE EFFECT OF JP-8 HYDROGEN CONTENT SPECIFICATION ON FINAL PRODUCT COSTS.

CASE NUMBER CASE NAME	1 Base	2 JP-8 13.4%	3 JP-8 13.0%	4 JP-8 13.0%
PLANT FLOWS, BPCD:				
Feeds:				
Butumen	25000	25000	25000	25000
Isobutane	6202	6202	6202	6202
Normal butane	2035	2174	2258	2284
Subtotal Feeds:	33237	33376	33460	33486
Products:				
Propane	980	977	976	976
Unleaded Gasoline	25117	26117	26736	26930
JP-4	0	0	0	0
JP-8	3782	2837	2269	2101
Residual Fuel	43	267	353	353
Subtotal Liquids	29922	30198	30334	30360
Sulfur, TPCD	51	51	51	51
Yields, Volume %:				
Total Liquids	90.0%	90.5%	90.7%	90.7%
Transportation Fuel	86.9%	86.8%	86.7%	86.7%
CAPITAL INVESTMENT, MM\$:				
ART	107.9	107.9	107.9	107.9
RCC	88.9	88.9	88.9	88.9
Recycle Hydrotreater	12.2	11.7	11.5	11.5
Naphtha Pretreater	0	0	0	0
JP-4 Hydrotreater	0	0	0	0
JP-8 Hydrotreater	16.2	13.5	12.4	12.4
Alkylation	14.5	14.5	14.5	14.5
Hydrogen Plant	12.6	11.4	10.9	10.9
Sulfur Plant	4.2	4.2	4.2	4.2
Battery Limits	256.5	252.1	250.3	250.3
Tankage	11.8	11.8	11.9	11.9
Offsites at 45%	115.5	113.5	112.7	112.7
Fixed Investment	383.8	377.4	374.9	374.9
OPERATING COSTS, MM\$/yr:				
Utilities	54.3	53.5	53.2	53.2
Fixed Costs	26.1	25.8	25.7	25.7
Byproduct Credits	-7.9	-9.5	-10.1	-10.1
Net Operating Costs	72.5	69.8	68.8	68.8
TRANSPORTATION FUEL COST COMPONENTS, \$/bbl:				
Startup	.23	.23	.23	.23
Working Capital	.29	.29	.29	.29
Byproducts	-.75	-.90	-.96	-.96
Fixed Costs	2.55	2.52	2.50	2.50
Income Taxes	3.77	3.68	3.67	3.67
Utilities	5.15	5.07	5.03	5.03
Debt Service	0.00	0.00	0.00	0.00
Capital	6.59	6.44	6.41	6.41
Raw Materials	26.00	26.09	26.12	26.13
Total Cost, \$/bbl	43.83	43.41	43.30	43.28
Incremental Turbine Fuel Cost over base, \$/bbl	0	39.54	37.09	36.36
Incremental Turbine Fuel	0	-945	-1513	-1681

The reduction in overall cost with reduced hydrogen specification is shown graphically in Figure 16. Of particular interest, by fixing gasoline value at the base level and allowing the value of JP-8 to float, JP-8 cost could actually fall as much as \$7.50/barrel.

In general, there is a strong driving force to reduce the required hydrogen content to about 13.3%. While obviously a potential problem in terms of smoke point and engine life, this reduction should be a representative target as more naphthenic fuel sources are investigated in future efforts.

Turbine Fuel Production Level

The United States military is critically interested in maximum potential turbine fuel supply for strategic reasons. As a result, several cases (Table 29) were evaluated to determine maximum JP-4, maximum JP-8, and maximum total turbine fuel levels. Predicted maximums were constrained at relatively low levels due to the blending streams used and hydrogen availability; yields approaching 70-80% would be feasible by moving the recycle hydrotreater between the ARTSM and RCCSM units.

Maximum predicted total turbine fuel production was about 14,000 BPD, or 45% of total feeds (57% based on bitumen). Increasing turbine fuel yields to this level reduced total plant production and increased product cost by \$1.30/barrel.

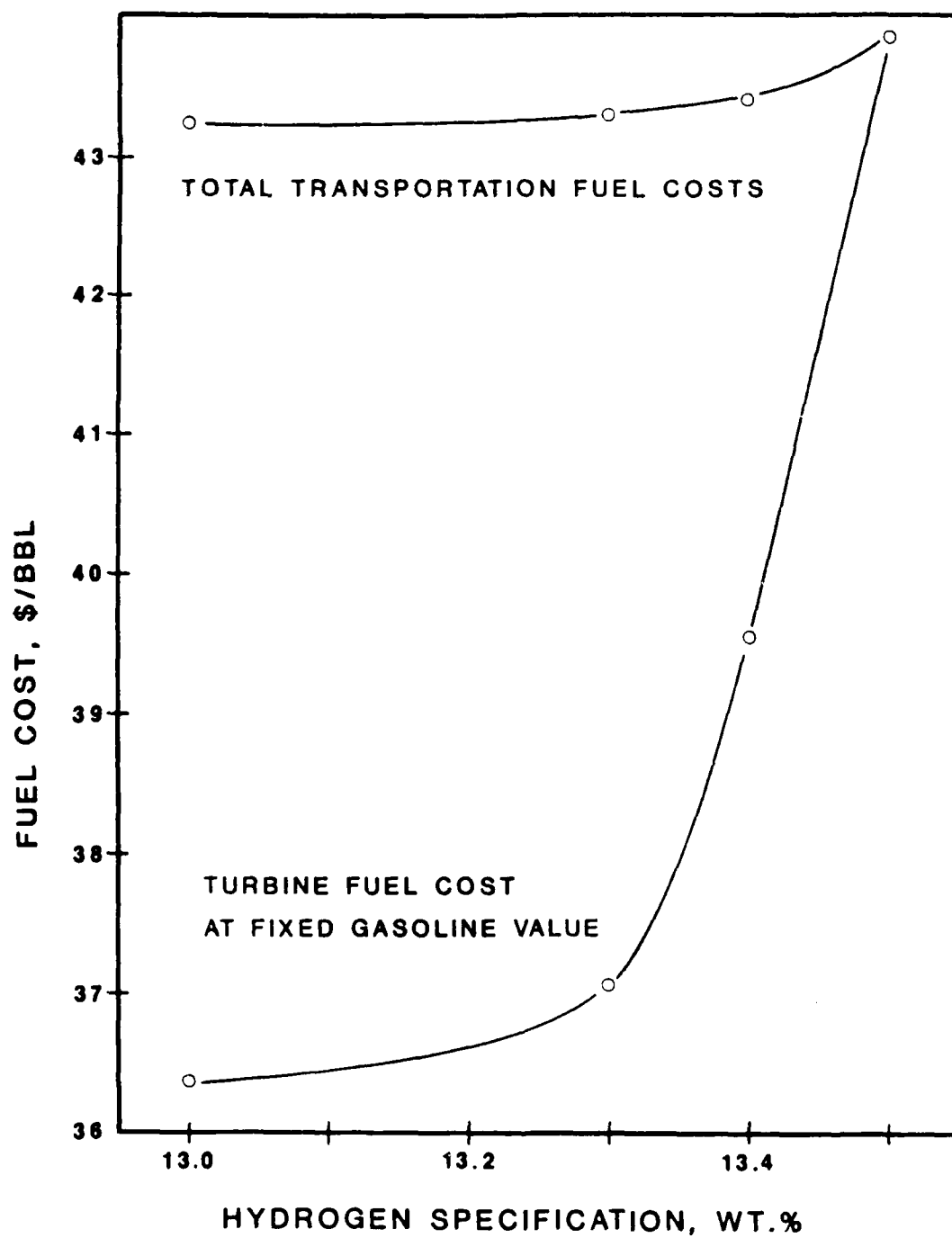


FIGURE 16. EFFECT OF HYDROGEN SPECIFICATION ON PRODUCT COSTS.

TABLE 29. COMPARISON OF THE COST EFFECT OF MAXIMIZING
TURBINE FUEL PRODUCTION LEVELS.

CASE NUMBER CASE NAME	1 Base	5 Max TF	6 Max JP-4	7 Max JP-8n	8 Max JP-8x
PLANT FLOWS, BPCD:					
Feeds:					
Butumen	25000	25000	25000	25000	25000
Isobutane	6202	5836	6202	6100	5836
Normal butane	2035	959	1026	1900	1912
Subtotal Feeds:	33237	31795	32228	33000	32748
Products:					
Propane	980	608	819	966	922
Unleaded Gasoline	25117	13146	14128	23903	23359
JP-4	0	12466	13164	0	0
JP-8	3782	1746	0	4704	4631
Residual Fuel	43	0	0	71	725
Subtotal Liquids	29922	27966	28111	29644	29637
Sulfur, TPCD	51	51	52	51	50
Yields, Volume %:					
Total Liquids	90.0%	88.0%	87.2%	89.8%	90.5%
Transportation Fuel	86.9%	86.0%	84.7%	86.7%	85.5%
CAPITAL INVESTMENT, MM\$:					
ART	107.9	108.6	107.9	108.1	108.6
RCC	88.9	80.2	88.9	86.5	80.2
Recycle Hydrotreater	12.2	11.6	12.7	12.3	11.3
Naphtha Pretreater	0	2.1	2.1	0	0
JP-4 Hydrotreater	0	10.4	10.9	0	0
JP-8 Hydrotreater	16.2	9.9	0	18.7	18.6
Alkylation	14.5	13.9	14.5	14.3	13.9
Hydrogen Plant	12.6	15.9	16	13.3	12.6
Sulfur Plant	4.2	4.2	4.3	4.2	4.2
Battery Limits	256.5	256.8	257.3	257.4	249.4
Tankage	11.8	11.4	11.3	11.7	11.8
Offsites at 45%	115.5	115.5	115.8	115.9	112.2
Fixed Investment	383.8	383.7	384.4	385	373.4
OPERATING COSTS, MM\$/yr:					
Utilities	54.3	55.4	56.0	54.7	53.7
Fixed Costs	26.1	26.7	26.1	26.1	25.6
Byproduct Credits	-7.9	-5.4	-6.7	-8.0	-12.5
Net Operating Costs	72.5	76.7	75.4	72.8	66.8
TRANSPORTATION FUEL COST COMPONENTS, \$/bbl:					
Startup	.23	.25	.25	.24	.24
Working Capital	.29	.29	.29	.29	.29
Byproducts	-.75	-.54	-.67	-.77	-1.22
Fixed Costs	2.55	2.76	2.70	2.58	2.58
Income Taxes	3.77	3.98	4.00	3.82	3.79
Utilities	5.15	5.55	5.62	5.24	5.26
Capital	6.59	6.96	6.98	6.68	6.62
Raw Materials	26.00	25.91	26.46	26.01	26.31
Total Cost, \$/bbl	43.83	45.14	45.63	44.09	43.86
Incremental Turbine Fuel Cost over base, \$/bbl	0	47.28	49.08	45.44	44.03
Incremental Turbine Fuel, BPCD:	0	10430	9382	922	849

Attributing the cost increase only to the incremental turbine fuel (10,430 barrels) produced an incremental fuel cost of \$47.30. This was a relatively low incremental cost for a large yield change.

Maximum JP-4 yields were slightly over 13,000 BPD. Compared to the base case, 9300 barrels of additional fuel were produced at an incremental cost of \$49.10/barrel, primarily due to higher utilities costs and lower total plant yield.

Two levels of JP-8 production were screened, using normal (JP-8n) and extended hydrogen and gravity (JP-8x) specifications. Less than 1000 barrels of additional JP-8 were produced in either case, but at very low (\$0.20-1.60/barrel) incremental cost. JP-8 production was very constrained by the flow scheme defined; significant increases should be available by hydro-treating ARTSM, rather than RCCSM, products.

Overall, the process was only moderately sensitive to varying turbine fuel production levels from zero to 14,000 barrels per day. The major change in this variation was increased hydrogen production and larger turbine fuel hydrotreaters.

SECTION VI

CONCLUSIONS

The overall program has shown, for a combination of ARTSM, RCCSM, and hydrotreating steps:

- This process shows excellent potential for production of high volumes of transportation fuels from bitumen and heavy oils. However, present conditions and crude availability make this option uneconomic in today's market.
- Excellent quality turbine fuels are available from this process. These fuels are naphthenic, with higher density than normal and with excellent thermal properties.
- The optimum process configuration requires hydrogen input to the conversion step for Westken, but not for higher native hydrogen content feeds such as Hondo. Hydrotreating between the ARTSM and RCCSM steps may be an improvement to the sequence.
- Desalting and diluent requirement reduction are keys to further cost reductions. Both are predicted to be attainable commercially, but were constrained by laboratory/pilot plant limitations.

- Cost reductions and higher density fuels are available by reducing the hydrogen content specification for JP-8 fuels. Future naphthenic fuel work should consider relaxation of the specification to 13.3% hydrogen.

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1. Moore, H. F., C. A. Johnson, W. A. Sutton, L. M. Henton, and M. H. Chaffin, "Aviation Turbine Fuels From Tar Sands Bitumen and Heavy Oils, Part I - Process Analysis," Contract F33615-83-C-2301, AFWAL-TR-84-2070, Part I, (September, 1984).
2. Moore, H. F., "Jet Fuel Production From Tar Sands and Heavy Oil by Asphalt Residual Treatment/Reduced Crude Conversion Process," USAF Aviation Turbine Fuels, 1985 Technology Review, Dayton, Ohio (March 26-27, 1985).
3. Moore, H. F., C. A. Johnson, D. A. Fabry, and M. H. Chaffin, "Aviation Turbine Fuels from Tar Sands Bitumen and Heavy Oils, Part II - Laboratory Sample Production," Contract F33615-83-C-2301, AFWAL-TR-84-2070, Part II, (July, 1987).
4. Busch, L. E., et. al., "Reduced Crude Oil Conversion in Commercial RCCSM and ARTSM Process Operations," 1984 NPRA Annual Meeting, San Antonio (March 25-27, 1984).

LIST OF SYMBOLS AND ABBREVIATIONS

°API	American Petroleum Institute liquid density scale
ART sm	Asphalt Residual Treatment, a service mark of Engelhard Corporation for professional services relating to selective vaporization processes for removing contaminants from petroleum feedstocks.
ASTM	American Society for Testing and Materials
BBL	barrels, 42 US gallons
BPCD	barrels per calendar day
BS&W	bottoms, sediment, and water
BPD	barrels per day
BTU	British Thermal Units
cc	cubic centimeter
CE	Chemical Engineering Magazine
cp	viscosity, centipoise
cs, cst	viscosity, centistokes
C ₃	propane
C ₄	butane
C ₅ ⁺	pentane and higher boiling hydrocarbons
D/B	diluent-to-bitumen ratio
DCF	Discounted cash flow
DF-2	diesel fuel
DOD	United States Department of Defense
DOE	United States Department of Energy
FIA	Hydrocarbon type analysis by fluorescent indicator adsorption

LIST OF SYMBOLS AND ABBREVIATIONS (CONT'D)

°F	temperature, degrees Fahrenheit
FCC	fluid catalytic cracker or cracking
FOE	fuel oil equivalent
gm	gram
Hg	mercury
Hr	hour
IBP	initial boiling point
IC ₄	isobutane
JP-4	MIL-T-5624L jet fuel
JP-8	MIL-T-83133A jet fuel
K factor	Watson K factor, defined as the cube root of the volumetric average boiling point, in °Rankine, divided by the specific gravity.
Kg	Kilogram
KwHr	Kilowatt-Hour
l	liter
lbs.	pounds, avoirdupois
LCO	light cycle oil
LHSV	liquid hourly space velocity
LP	linear programming
M	thousand
m	meter
MM	million

LIST OF SYMBOLS AND ABBREVIATIONS (CONT'D)

mm	millimeter
m ³	cubic meter
MAT	microactivity test
max.	maximum
mg	milligram
min	minimum
ml	milliliter
N _B	basic nitrogen content
NC ₄	normal butane
Ni	nickel
No.	number
N _T	total nitrogen content
OP.	operation
pH	negative logarithm of hydrogen ion concentration
ppm	part per million (by weight unless specified)
psig	pounds per square inch gauge pressure
RCC sm	Reduced Crude Conversion, a registered service mark of Ashland Oil, Inc., for technical assistance and consulting services in connection with processes for heavy oil cracking and related catalysts.
RONC	research octane number, clear
RVP	Reid vapor pressure, psig
SCFB	standard cubic feet per barrel (42 gallons)
Sim-D	Simulated Distillations by Gas Chromatography
SUS	Viscosity, Saybolt Universal Seconds

LIST OF SYMBOLS AND ABBREVIATIONS (CONT'D)

TBP	True Boiling Point
Tot	Total
TPD	U. S. tons per day
Trans.	Transportation
Typ	typical
USAF	United States Air Force
V	vanadium
vol	volume
WHSV	weight hourly space velocity
wt	weight
<	less than
>	greater than
@	at
%	percent
°	degrees
()	byproduct credits when used in economic value tables
"	inch
\$	US dollars

APPENDIX A

DETAILED SAMPLE PREPARATION FLOWS

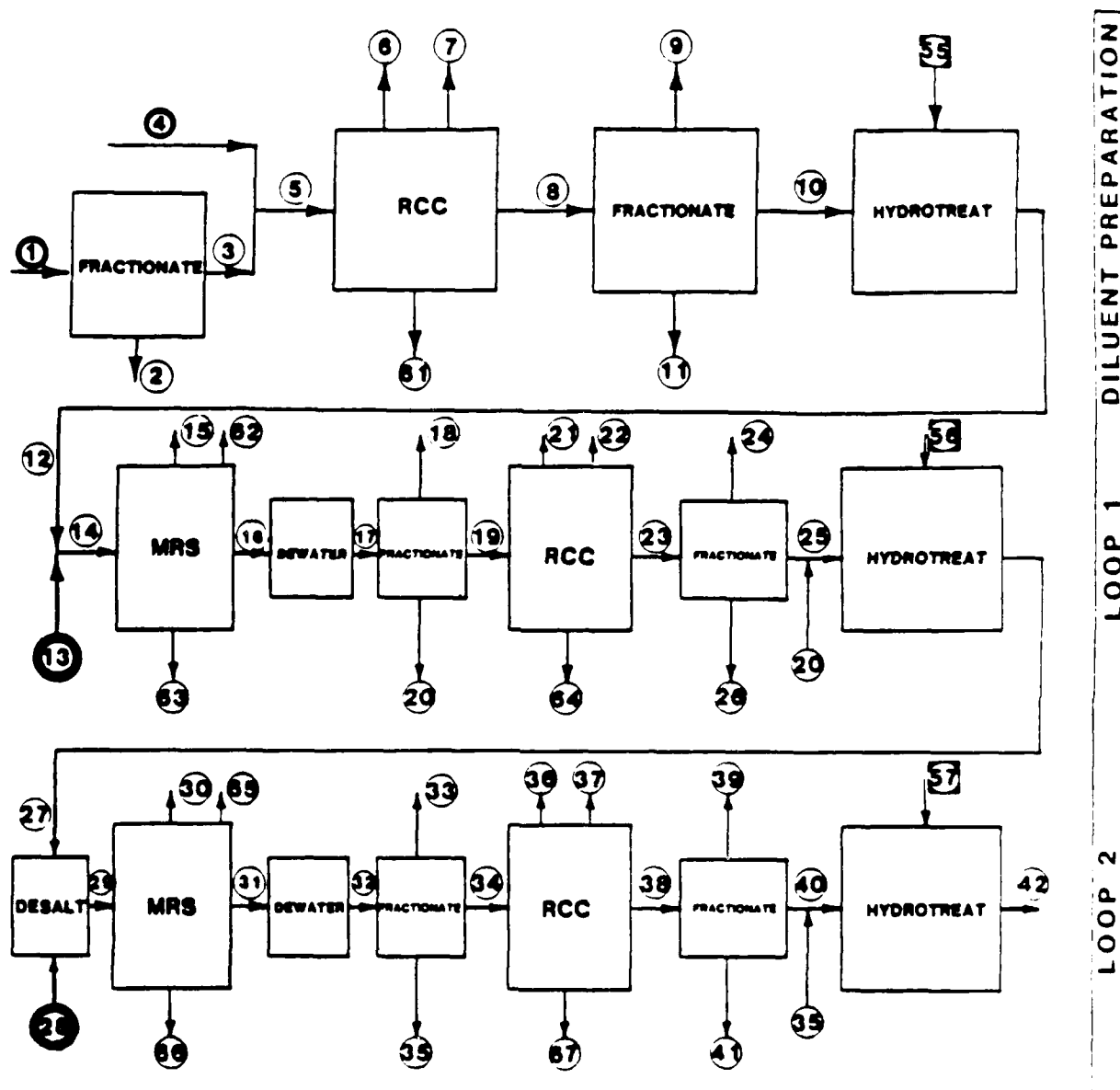


FIGURE A-1. SAMPLE PREPARATION CONVERSION SECTION

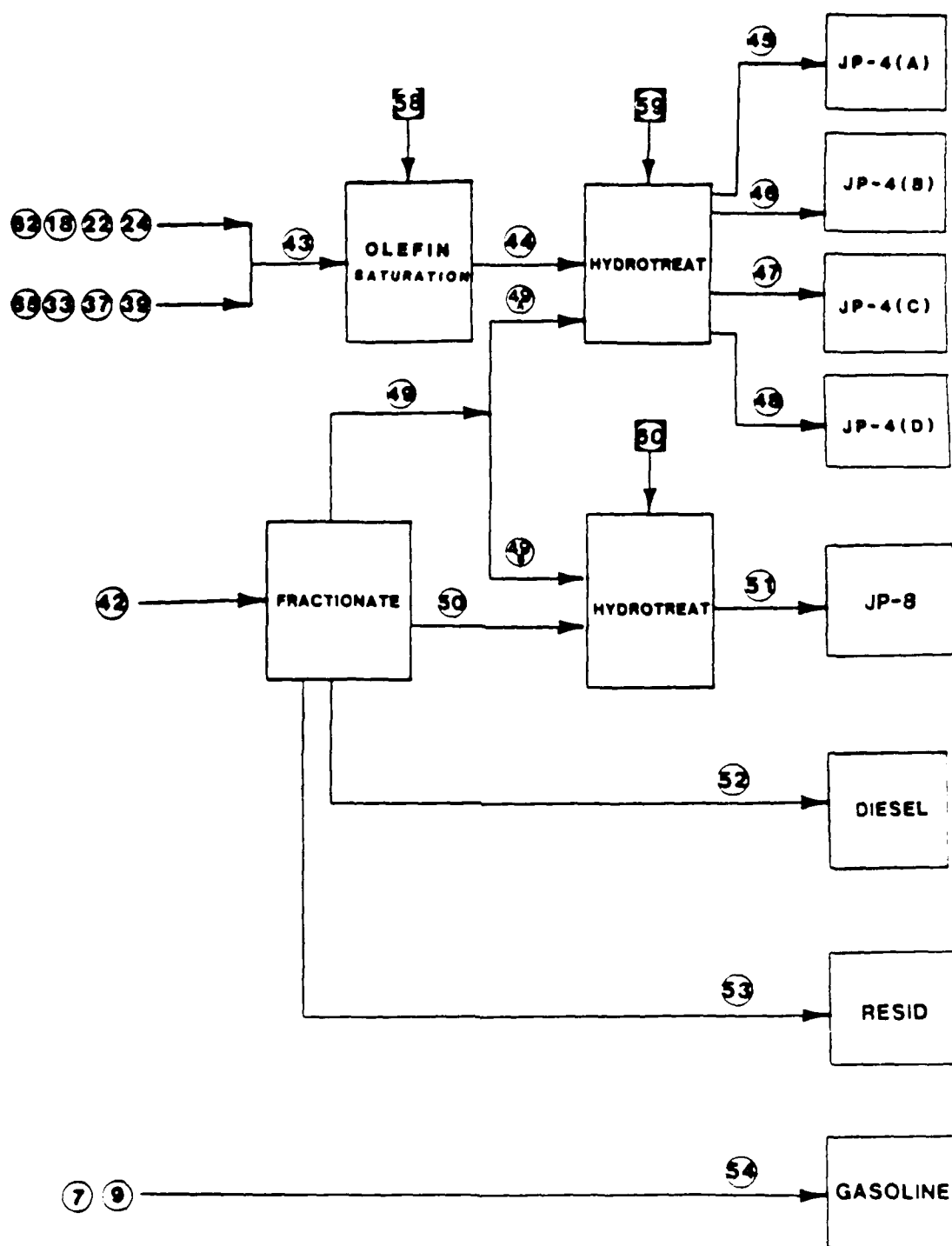


FIGURE A-4. SAMPLE PREPARATION FINAL FUEL HYDROTREATING

TABLE A-1. PHASE III SAMPLE PREPARATION ACTUAL MATERIAL FLOWS

STREAM IDENTIFICATION	QUANTITY, lbs	COMMENTS
1. RAW WESTKIN BITUMEN	2389	
2. WFE RESID (1000°F-)	1214	
3. WFE GAS OIL (1000°F-)	1175	
4. RAW WESTKIN BITUMEN	800	
5. COMBINED GAS OIL BITUMEN FEED	19.6	
6. RCC OFF GAS	229	
7. RCC COLD TRAP LIQUIDS	77	
8. RCC FULL RANGE LIQUID PRODUCT	1229	
9. RCC NAPHTHA (IBP-330°F)	228	
10. RCC CYCLE OIL (330-800°F)	930	
11. RCC RESID (800°F-)	0	
12. HYDROTREATED CYCLE OIL DILUENT	890	
13. RAW WESTKIN BITUMEN	51	
14. COMBINED BITUMEN CYCLE OIL DILUENT	1106	
15. WRS OFF GAS	63	
16. WRS LIQUID PRODUCT	860	
17. DEWATERED WRS PRODUCT	870	
18. WRS LIGHT NAPHTHA (IBP-330°F)	64	
19. WRS BOTTOMS (430°F-)	655	
20. WRS HEAVY NAPHTHA (330-430°F)	160	
21. RCC OFF GAS	78	
22. RCC COLD TRAP LIQUIDS	45	
23. RCC LIQUID PRODUCT	471	
24. RCC NAPHTHA (IBP-330°F)	82	
25. RCC CYCLE OIL (330-800°F)	388	
26. RCC RESID (800°F-)	0	
27. HYDROTREATED CYCLE OIL (LOOP 1)	487	
28. RAW WESTKIN BITUMEN	650	
29. DESALTED BITUMEN CYCLE OIL	1031	
30. WRS OFF GAS	59	
31. WRS LIQUID PRODUCT	809	
32. DEWATERED WRS PRODUCT	740	
33. WRS LIGHT NAPHTHA (IBP-330°F)	36	
34. WRS BOTTOMS (430°F-)	527	
35. WRS HEAVY NAPHTHA (330-430°F)	119	
36. RCC OFF GAS	59	
37. RCC COLD TRAP LIQUIDS	43	
38. RCC LIQUID PRODUCT	402	
39. RCC NAPHTHA (IBP-330°F)	67	
40. RCC CYCLE OIL (330-800°F)	313	
41. RCC RESID (800°F-)	0	
42. HYDROTREATED CYCLE OIL (LOOP 2)	403	
43. COMBINED NAPHTHA	260	
44. OLEFIN SATURATED NAPHTHA	193	
45. JP-4(A), 35% AROMATICS	83	
46. JP-4(B), 30% AROMATICS	63	
47. JP-4(C), 25% AROMATICS	33	
48. JP-4(D), ON SPECIFICATION	71	
49. IBP-540°F CYCLE OIL (TOTAL)	129	
49A. IBP-540°F CYCLE OIL (PART A)	80	
49B. IBP-540°F CYCLE OIL (PART B)	47	
50. 540-640°F CYCLE OIL	111	
51. SPECIFICATION JP-8	78	
52. DIESEL		
53. RESID		
54. BLENDED GASOLINE		
55. HYDROGEN	8.6	
56. HYDROGEN	4.6	
57. HYDROGEN	5.9	
58. HYDROGEN	0.9	
59. HYDROGEN	1.8	
60. HYDROGEN	1	
61. RCC COKE	329	
62. WRS COLD TRAP LIQUIDS	9	
63. WRS COKE	125	
64. RCC COKE	36	
65. WRS COLD TRAP	3	
66. WRS COKE	155	
67. RCC COKE	33	

APPENDIX B

DETAILED ECONOMIC CASE STUDY FLOW DIAGRAMS

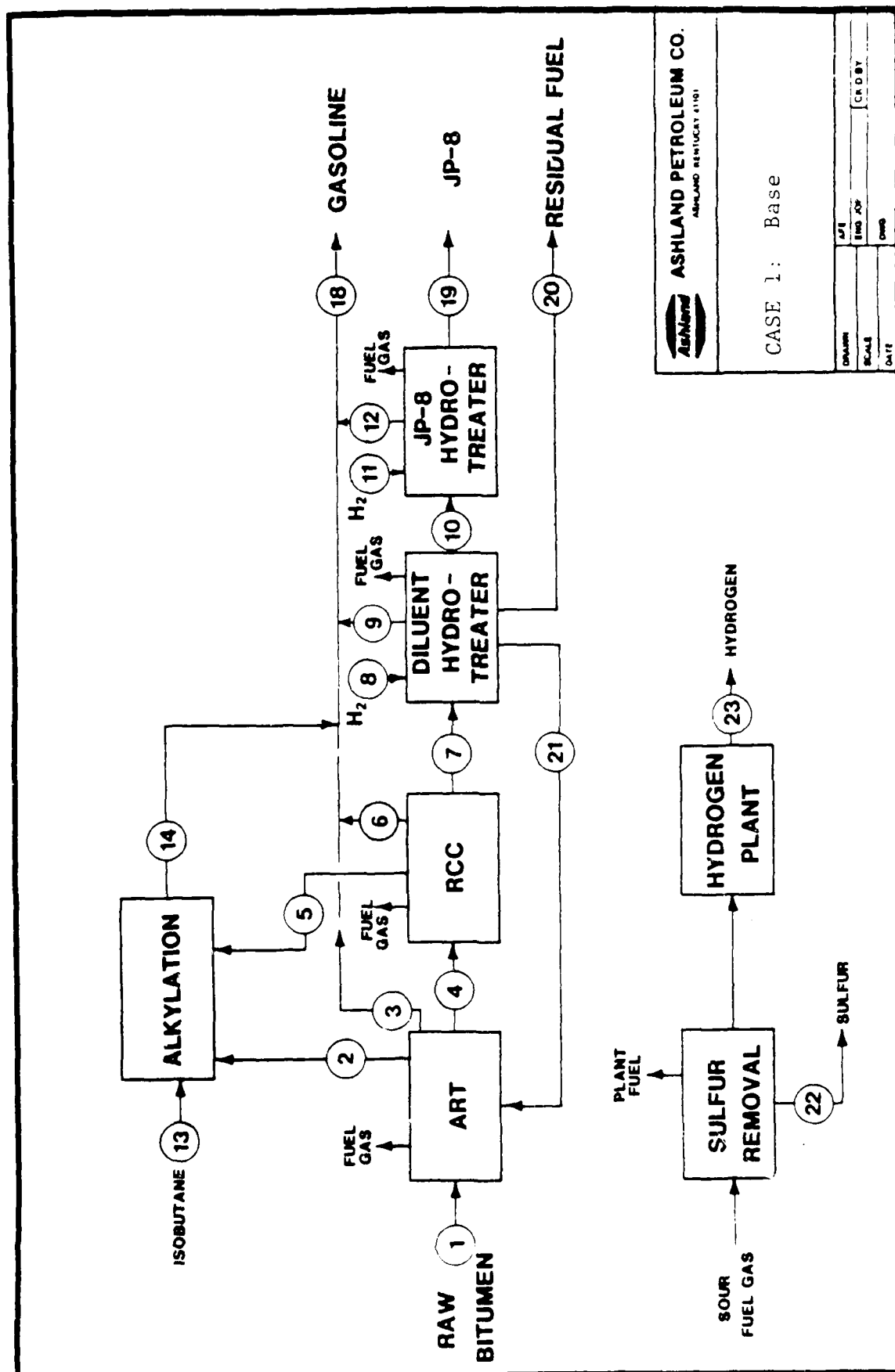


FIGURE B-1. BASE CASE PLANT FLOWS

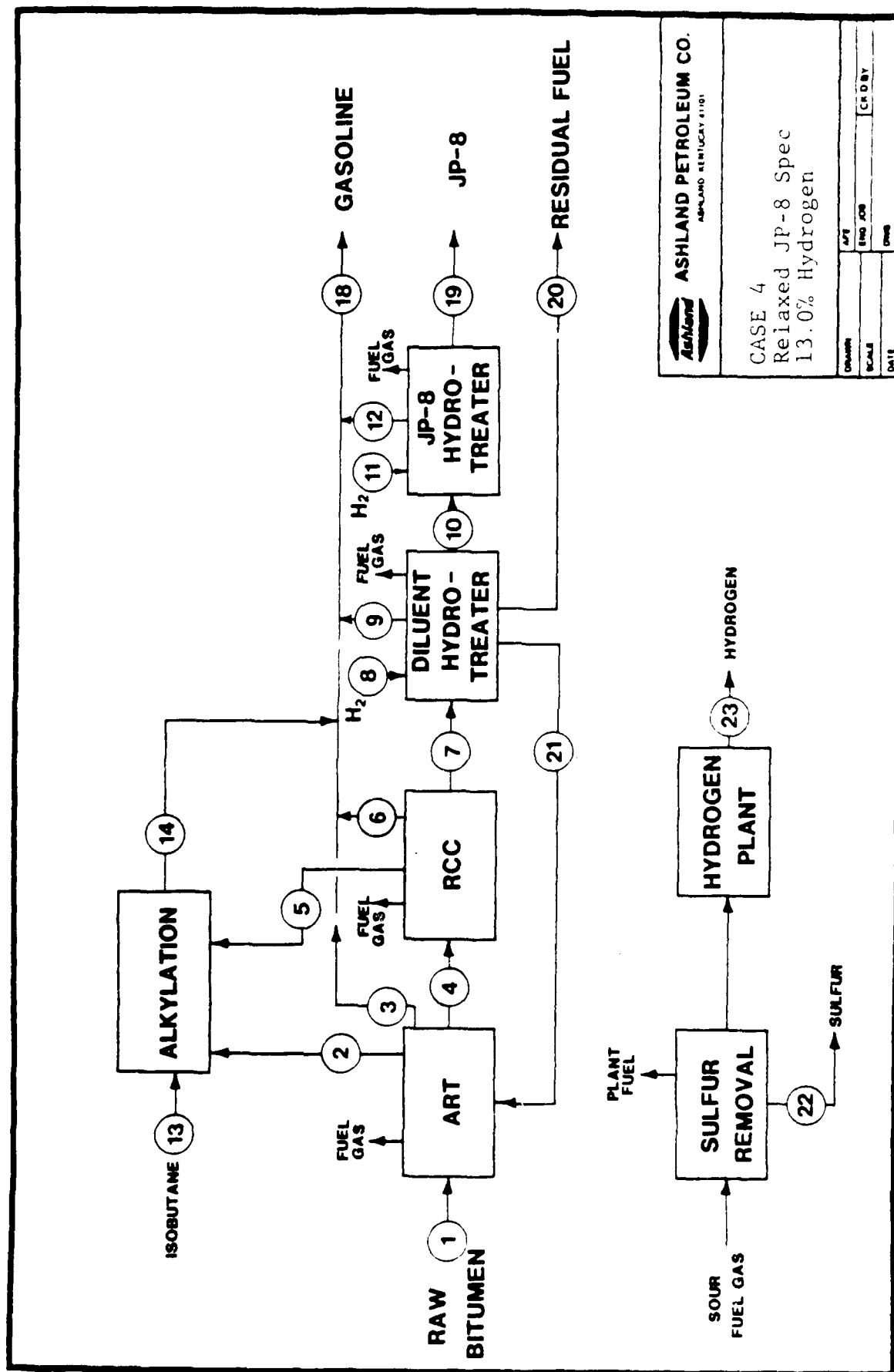


FIGURE B-2. LOWERED JP-8 HYDROGEN CONTENT



FIGURE B-3. MAXIMUM TURBINE FUEL PRODUCTION LEVELS

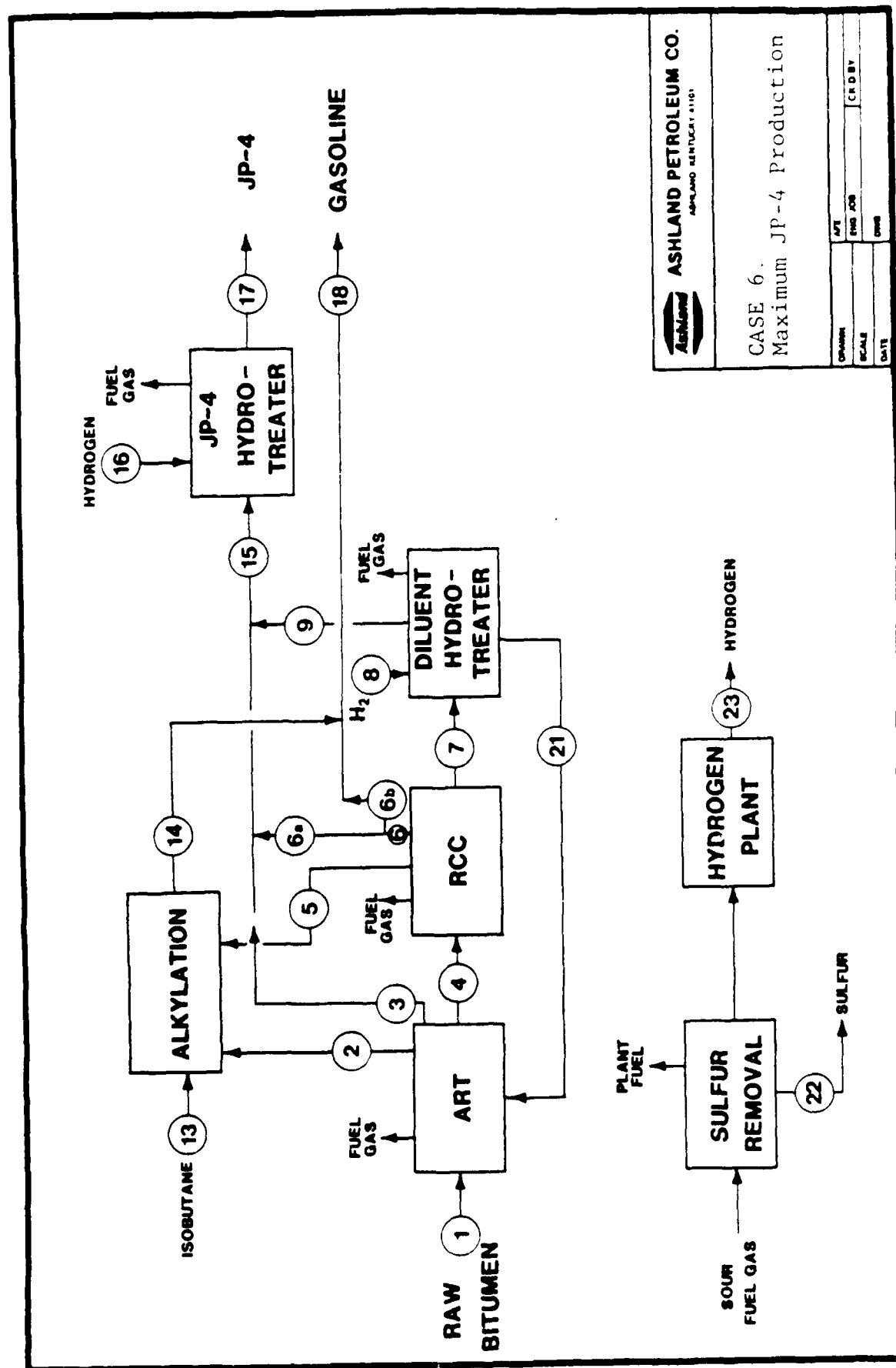
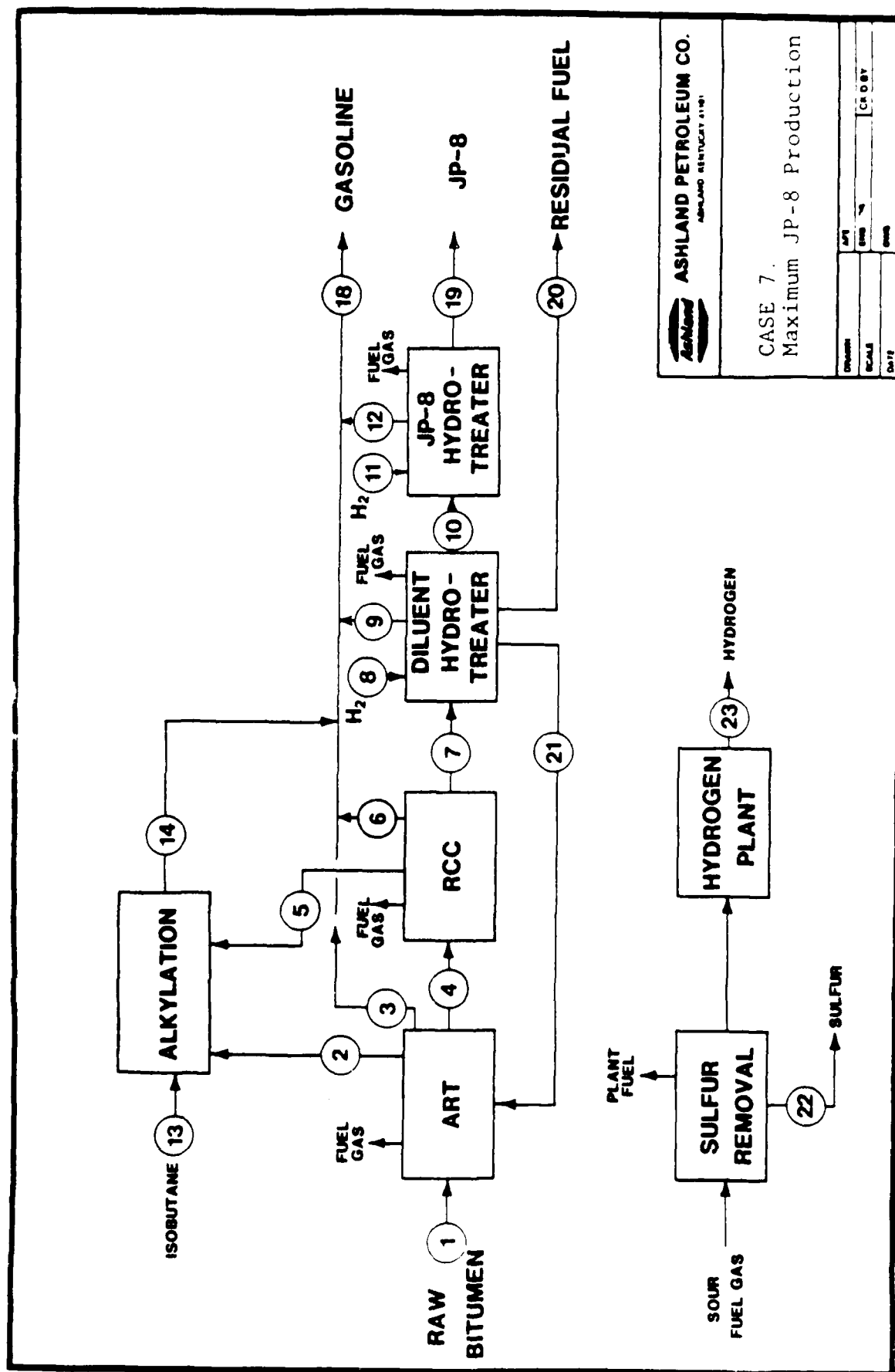


FIGURE B-4. MAXIMUM JP-4 PRODUCTION LEVELS

 ASHLAND PETROLEUM CO. ASHLAND KENTUCKY 41101			
CASE 6. Maximum JP-4 Production			
DRAWN DATE	DESIGNED DATE	CHECKED DATE	CREDIT DATE



 ASHLAND PETROLEUM CO. ASHLAND KENTUCKY 41101			
CASE 7. Maximum JP-8 Production			
Drawn _____	Date _____	Scale _____	Job No. _____
Rev. _____	Date _____	Scale _____	Job No. _____

FIGURE B-5. MAXIMUM JP-8 PRODUCTION LEVELS

TABLE B-1

DETAILED FLOW DEFINITION

ALL UNITS ARE TONS PER CALENDAR DAY

<u>Stream Number</u>	<u>Identification</u>	<u>Case 1</u>	<u>Case 2</u>	<u>Case 5</u>	<u>Case 6</u>	<u>Case 7</u>
1.	Raw Tar Sands Bitumen	4363	4363	4363	4363	4363
2.	ART sm C ₃ +C ₄	110	110	110	110	110
3.	ART sm Naphtha	689	689	829	689	728
4.	ART sm Product, 430°F+	4600	4600	3883	4600	4400
5.	RCC sm C ₃ +C ₄	417	417	379	417	406
6.	RCC sm Naphtha, C ₅ -430°F	1153	1348	840	1271	964
6a.	RCC sm Naphtha to JP-4 Hydrotreater	--	--	344	739	--
6b.	RCC sm Naphtha to Gasoline Pool	--	--	496	532	--
7.	RCC sm Product, 330°F+	2562	2367	2251	2704	2579
8.	Hydrogen	45	40	40	48	46
9.	Hydrotreated Diluent, C ₅ -330°F	19	12	488	355	23
10.	Hydrotreated Diluent, 330-430°F	595	403	287	--	740
11.	Hydrogen	18	12	9	--	23
12.	Hydrotreated C ₅ -330°F	52	44	30	--	65
13.	Isobutane	630	630	586	630	618
14.	C ₃ +C ₄ Alkylate	1077	1077	1002	1077	1056
15.	Combined Naphtha	--	--	1661	1783	--
16.	Hydrogen	--	--	38	39	--
17.	Finished JP-4	--	--	1720	1814	--
18.	Finished Gasoline	3028	3222	1498	1609	2836
19.	Finished JP-8	554	315	263	--	690
20.	Residual Fuel	140	140	--	--	174
21.	Hydrotreated Diluent, 430°F+	2340	2340	2216	2340	2305
22.	Sulfur	51	51	51	52	51
23.	Hydrogen (100% Basis)	64	53	87	88	69

TABLE B-2

NOMINAL PROCESS CONDITIONS

BASE CASE SOLUTION

<u>Unit</u>	<u>Nominal Temperature, °F</u>	<u>Pressure, PSIG</u>	<u>Catalyst Ratio</u>	<u>Hydrogen Circulation, SCFB</u>
ART	940	10	4	None
RCC	960	10	8	None
Diluent Hydrotreater	700	1400	1.5	3000
JP-8 Hydrotreater	690	2000	0.5	4000

Notes: A - Weight of catalyst circulated per weight of oil feed.

B - LHSV, volume of oil feed (as liquid at 60°F) per volume
of catalyst per hour

END
DATE
FILMED
MARCH
1988
DTIC