No. 28

OF

THE STUDY REPORT
ON
THE ORINOCO HEAVY OIL UPGRADING PROJECT
FOR
THE REPUBLIC OF VENEZUELA

VOLUME I : REPORT

NOVEMBER 1980

JAPAN INTERNATIONAL COOPERATION AGENCY

PREFACE

It is with great pleasure that I present this report entitled "The Report on the Orinoco Heavy Oil Upgrading Project for the Republic of Venezuela" to the Government of the Republic of Venezuela.

This report embodies the results of the first survey and the second survey which were carried out (in Caracas and Orinoco area, Venezuela) from September 30 to October 13, 1979 and from May 3 to May 23, 1980 by the Japanese survey team commissioned by the Japan International Cooperation Agency following the request of the Government of the Republic of Venezuela.

The survey team, headed by Mr. Senichi Hirose, had a series of close discussions with the officials concerned of the Government of the Republic of Venezuela and conducted a wide scope of field survey and data analyses.

I sincerely hope that this report will be useful as a basic reference for development of the project.

I am particularly pleased to express my appreciation to the officials concerned of the Government of the Republic of Venezuela for their close cooperation extended to the Japanese team.

November 1980

Keisuke Arita
President
JAPAN INTERNATIONAL COOPERATION AGENCY

CONTENTS

GLOSSARY

°API	gravity of petroleum defined by American Petroleum Institute
BBL	barrel
BPCD	barrel per calendar day
BPSD	barrel per stream day
BTU	British Thermal Unit
°C	degree centigrade
cp	centipoise
cst	centistokes
DAO	deasphalted oil
DCF	discounted cash flow
°F	degree fahrenheit
FOE	The heating value of a standard barrel of crude oil, equal to 6.24 million BTU (LHV)
Hr or H	hour
HCR	hydrocracking (unit)
HDS	hydrodesulfurization (unit)
HTR	hydrotreating (unit)
HP	high pressure
H_2 Plant	hydrogen generation plant
INTEVEP	Instituto Technológico Venezolano del Petróleo
JICA	Japan International Cooperation Agency
Kcal	kilocalorie
Kg	kilogram
LAGOVEN	Lagoven S.A.
LP	low pressure
MEM	Ministerio de Energía y Minas
m	meter
m^2	square meter
m^3	cubic meter
MM	million
MMBTU	million British Thermal Unit
MMKcal	million kilocalorie
$MMNm^3/SD$	million normal cubic meter per stream day
MMSCFD	million standard cubic feet per day
MW	megawatt
MP	medium pressure
m^3/H	cubic meter per hour

Nm^3	normal cubic meter
MH	man-hour
MT	metric ton
%	percent
wt.%	weight percent
vol.%	volume percent
PDVSA	Petróleos de Venezuela, S.A.
ppm	parts per million
ROE	rate of return on equity
ROR	rate of return
SCF	standard cubic feet
SDA	solvent deasphalting (unit)
SCFD	standard cubic feet per day
Sp.Gr.	specific gravity
SR	sulfur recovery (unit)
SCF/B	standard cubic feet per barrel
Ton/H or T/H	tons per hour
Ton/SD or T/SD	tons per stream day
Ton/Y or T/Y	tons per year
UHP	Ultra high pressure
US$	US dollar
10^6 US$	million US dollar
US$/BBL	US dollar per barrel
US$/MMBTU	US dollar per million BTU
y	year
wt.%S	weight percent sulfur
VGO	vacuum gas oil

FIGURES

TABLES

CHAPTER 1 INTRODUCTION

Volume I is a report dealing with the technical and economic studies on the Orinoco heavy oil upgrading project. With a view to comparing and evaluating the three upgrading processes, the study in this report plans on three cases of upgrading refineries based on these processes, and weighes with one another on the entire refineries including process units, utilities, and offsite facilities.

Volume I has been compiled using a separate Volume II, "Supplement" as the data base. Chapters 1, 2 and 3 of Volume II give three reports submitted by the three respective groups who have proposed their own processes in response to the same prerequisite conditions.

CHAPTER 2 SUMMARY

2.1 OBJECTIVES OF STUDY

The study is related to the three processes proposed by three groups of Japan, which are the Toa Oil Co., Ltd. group, Kureha Chemical Industry Co., Ltd. group and Maruzen Oil Co., Ltd. group, for the upgrading of the heavy crude oil to be produced under the Orinoco Heavy Crude Development Project in an area of 42,000 km^2 wide located on the north side of the River Orinoco in the Republic of Venezuela.

The objectives of the study are:

- to make clear the respective features of the three processes and
- to provide information for the process selection in the course of planning a commercial plant for the upgrading of the heavy crude oil

2.2 UPGRADING REFINERY SITE

The Orinoco Oil Belt lies on the north side of the River Orinoco, and the area has a width of 70 km and a length of 600 km which covers the delta of the river, Monagas state and Anzoategui state and the south of Guarico state.

Fig. 2.1 shows a general map of the Orinoco area.

2.3 PROJECT OUTLINE

The upgrading schemes of the heavy crude oil proposed by three groups of Japan are as follows:

- scheme based on the Fluid Coking process proposed by Toa Oil Co., Ltd. group (hereinafter referred to as "Fluid Coker Case")
- scheme based on the Eureka process proposed by Kureha Chemical Industry Co., Ltd. group (hereinafter referred to as "Eureka Case")
- scheme based on the M-DS process proposed by Maruzen Oil Co., Ltd. group (hereinafter referred to as "M-DS Case")

2.3.1 Crude Oil and Product

50/50% Cogollar IX and Cerro Negro Crude Oil is processed to yield 125,000 BPSD improved crude oil having a gravity of 25–28° API and sulfur content of less than 1.0 wt. percent.

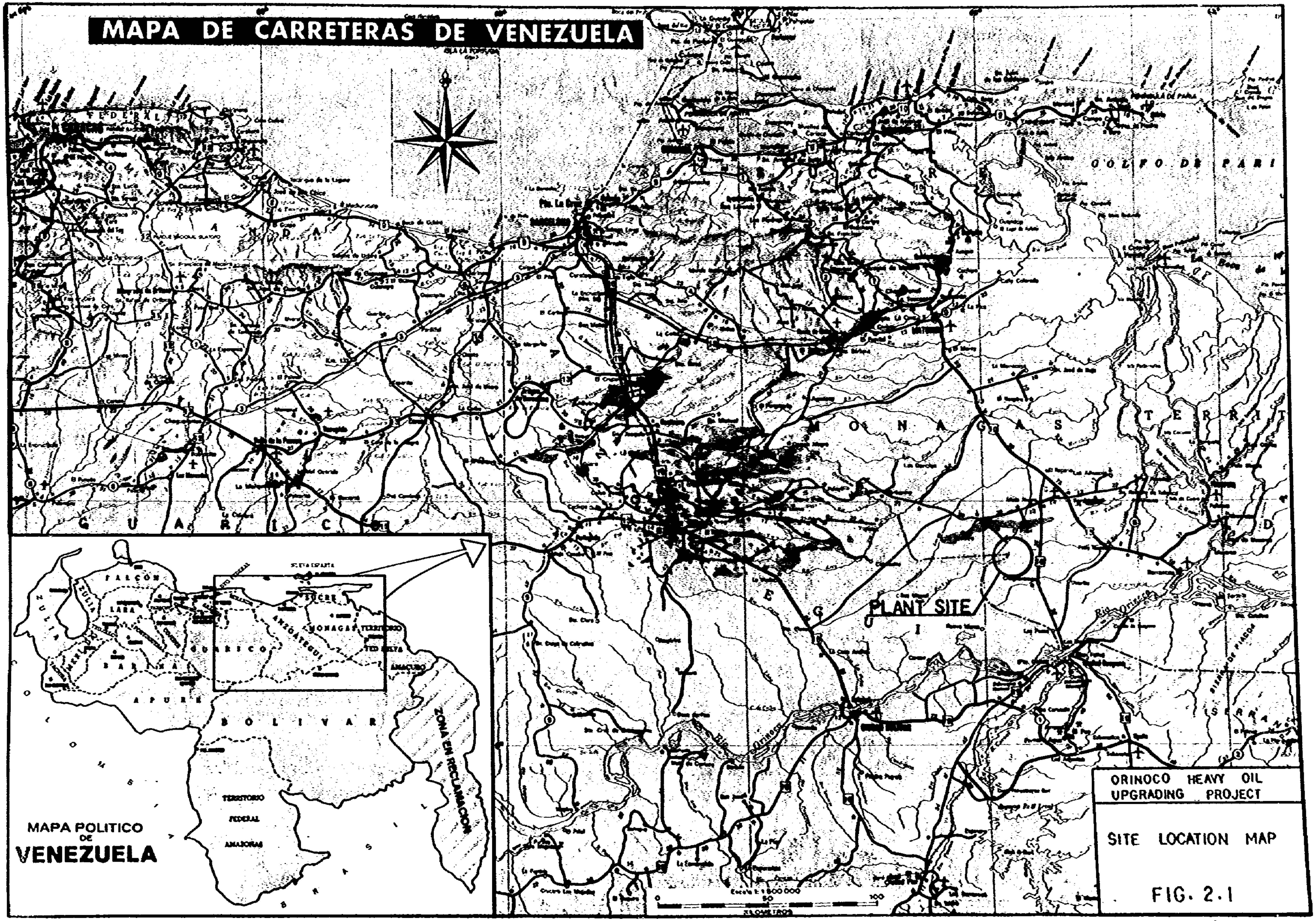
Overall material balances and the characteristics of the improved crude oils are shown in Table 2.1 and Table 2.2, respectively.

Table 2.1 Overall Material Balance

Normal operation case

Case	Fluid Coker	Eureka	M-DS
Feed			
Raw Crude Oil	158,160	158,710	151,055
Diluent Light Gas Oil	47,428	47,613	45,317
Mixed Crude Oil, BPSD	205,588	206,323	196,372
Natural Gas, $MMNm^3/SD$	0.198	0.518	0.850
Products			
Improved Crude Oil, BPSD (Synthetic Crude Oil)	125,000	125,000	125,000
Sulfur, T/SD	509.2	569.0	558.0
Diluent Light Gas Oil, BPSD	47,428	47,613	45,317
Gypsum, T/SD*	673	498	594
Fuel Gas to Field $MMNm^3/SD$	1.08	0	0
Electric Power MW*	126.2	126.6	120.5
Surplus By-product Fuel, T/Y	746,345	899,992	792,126

* 365 D/Y production



ORINOCO HEAVY OIL UPGRADING PROJECT

SITE LOCATION MAP

FIG. 2.1

Table 2.2 Properties of Improved Crude Oils

	Fluid Coker	Eureka	M-DS
Feed			
(Raw Crude Oil)			
°API	8.5	8.5	8.5
Sulfur, wt%	3.67	3.67	3.67
Products			
(Improved Crude Oil)			
°API	25.7	25.0	26.1
Sulfur, wt.%	0.70	0.41	0.05
Viscosity			
cst. @ 100°F	4.9	–	22.0
cst. @ 210°F	2.8	–	4.0
Nitrogen. wt.%	0.17	–	0.008
CCR, wt.%	0.147	–	0.13
Components, vol. %			
SR. Naphtha	–	1.0	0.2
HTR. SR. Naphtha	1.0	–	–
HTR. SR. LGO	5.5	5.3	5.1
HTR. SR. HGO	12.9	12.6	12.1
HTR. VGO	35.9	35.1	35.4
HTR. Coker Naphtha	11.2	–	–
HTR. Coker Gas Oil	33.5	–	–
HTR. Cracked Light Oil	–	7.8	–
HTR. Cracked Heavy Oil	–	38.2	–
HTR DAO	–	–	47.2
Yield of Distillation			
C_5/375°F, vol. %	15.0	7.3	9.5
375/650°F, vol. %	30.0	32.4	34.0
650/1,000°F, vol. %	55.0	60.3	33.5
1,000°F+, vol. %	–	–	23.0
Sulfur content of distillation			
C_5/375°F, wt. %	0.24	0.09	0.01
375/650°F, wt. %	0.67	0.1	0.08
650/1,000°F, wt. %	0.73	0.6	0.02
1,000°F+, wt. %	–	–	0.03

2.3.2 Major Facilities

Three process schemes consist of the primary upgrading processes such as the Fluid Coking process, the Eureka process and the M-DS process, the pretreating processes and such secondary upgrading processes as hydrodesulfurization, sulfur recovery process and so on.

Utility facilities are planned on the basis of self-sufficiency.

Such offsite facilities as the crude oil storage, the improved oil storage and loading are planned based on the specific conditions of the project.

Major facilities planned for the respective three cases are shown in Table 2.3.

Table 2.3 Major Facilities

		Fluid Coker	Eureka	M-DS
Process Units				
Atmos. Crude Distillation	BPSD	102,800x2	103,200x2	98,200x2
Vacuum Flashing	BPSD	67,200x2	67,400x2	64,200x2
Fluid Coker	BPSD	43,600x2	–	–
Eureka	BPSD	–	42,400x2	–
M-DS	BPSD	–	–	40,700x2
HTR or HDS	BPSD	60,900x2	16,100x2	10,700x2
			45,900x2	48,800x2
Hydrogen Generation	$MMNm^3/SD$	0.90x2	1.08x2	1.70x2
Acid Gas Treating	Ton/SD-H_2S	68x2	315x2	309x2
		23x2		
		192x2		
Sulfur Recovery	Ton/SD-S	255x2	285x2	279x2
Utility Systems				
Steam				
Generator 100kg/cm²G	Ton/H	260x(3+1)	240x(5+1)	240x(5+1)
50kg/cm²G	Ton/H	200x2		
Power Generator	KW	55,000x(3+1)	46,000x(5+1)	44,000x(5+1)
		18,000x2		
BFW Treating	Ton/H	250x(2+1)	310x(2+1)	200x(2+1)
Cooling Water System	Ton/H	18,000x2	20,000x2	15,500x2
Tankage System				
Total Tank Capacity	10^3 Kl	1,436	1,397	1,340.5

2.4 ECONOMIC STUDY

2.4.1 Capital and Operating Costs

For the three case schemes, the capital and operating costs are estimated. In the estimation, all costs expressed in US dollars are based on the 1980 prices (escalation factor is not included) and reflect conditions in Venezuela.

Table 2.4 shows a summary of costs.

Table 2.4 Cost Summary

Case	Capital Requirement 10^6 US$	Direct Operating Cost US$/BBL of Improved Crude
Fluid Coker	1,073.40	1.63
Eureka	1,097.50	1.62
M-DS	1,188.18	2.15

The above costs are for process units and facilities inside the battery limits of the refinery. The cost for the following facilities are not included:

- Raw crude oil processing facility
- Raw material supply pipeline and facility
- Improved crude oil pipeline
- Raw water and natural gas supply pipeline
- Housing for the refinery's employees
- Transmission facilities for diluent, fuel gas and electric power
- Transmission facilities for sulfur, by-product and gypsum
- Storage facilities of surplus by-product

2.4.2 Economic Analysis

In order to evaluate the three cases from an economic viewpoint, an economic analysis is conducted by calculating the internal rate of return (IRR) based on assumed raw crude oil cost (US$10/BBL) and improved crude oil price (Tia Juana Medium crude price is adjusted by sulfur and gravity differentials). And sensitivity analyses are conducted by taking crude oil cost, construction cost and income tax as parameters. The details of the bases are described in section 3.8. The results of study are shown in Tables 2.5.

Table 2.5 Internal Rate of Return

	CASE	Fluid Coker	Eureka	M-DS
			ROE, %	
Income tax	50%			
Construction cost	Base	25.0	22.9	23.1
Raw crude oil	Base			
Income tax	67%			
Construction cost	Base	18.7	17.1	17.2
Raw crude oil	Base			

2.5 OVERALL COMPARISON

In this section, overall comparison and review are given on the three upgrading schemes proposed by three groups. The results of study will be described in Chapter 5 through Chapter 10. Major items having some differences in results among three schemes are enumerated in Table 2.6 "Comparison of Three Upgrading Schemes".

Details of comparison are described in Chapter 11.

2.5.1 Comparison from a Technical Viewpoint

Major differences exist in process schemes (combinations of processes), product, operation and combustion of by-product.

(1) Process scheme and product

(a) The yield of improved crude oil derived from the raw crude oil on a volume basis is highest in the M-DS case. And the properties of improved crude oil in the M-DS case are close to the target quality. This is because of the following: Unlike the Eureka and Fluid coker processes which are cracking processes, the M-DS process is a solvent deasphalting process which can not yield light oil sufficient for use in the improved crude oil of 25°API; therefore a direct hydrodesulfurization process of high severity is adopted for the purposes of desulfurization and hydrocracking in the M-DS case. For this reason, the requirement of natural gas in the M-DS case is greater than in other cases.

(b) Only Fluid coker case can supply off gas to the raw crude oil production field, because a lot of gas is produced by fluid coker.

(2) Operation

(a) Fluid coker and M-DS processes are operated continuously, but the Eureka process is operated on a semi-batchwise basis.

Table 2.6 Comparison of Three Upgrading Schemes

Item / Case	Fluid Coker		Eureka		M-DS	
Process Scheme and Yield						
Properties of improved crude oil						
°API	25.7		25.0		26.1	
Yield and Sulfur content	vol. %	wt. % S	vol. %	wt. % S	vol. %	wt. % S
C_5/375°F	15	0.24	7.3	0.09	9.5	0.01
375/650°F	30	0.67	32.4	0.1	34.0	0.08
650/1,000°F	55	0.73	60.3	0.6	33.5	0.02
1,000°F+	–	–	–	–	23.0	0.03
Yield of improved oil, vol. % on raw crude oil	79.0		78.8		82.8	
Natural gas requirements	0.198 MMNm³/D		0.518 MMNm³/D		0.850 MMNm³/D	
Off gas to crude oil field	1.08 MMNm³/D					
Upgrading process	Coking		Thermal cracking		Solvent deasphalting	
Hydrotreating and hydro-desulfurization process (HTR and HDS)	Hydrotreating of of cracked oil		Hydrotreating of of cracked oil		Hydrodesulfurization and hydrocracking	
Operation						
Upgrading process (Fluid Coker, Eureka, M-DS)	Continuous		Semi-batchwise High temperature and high viscosity pitch is rundown from pitch drum and sent to pitch flaker		Continuous High temperature and high viscosity asphalt is sent directly to boiler	
Commercial installation	More than 10		One is in operation. One is under construction		Many similar units	
HTR or HDS						
H_2 Consumption, SCF/B	HTR: 550		No. 1 HTR 440 No. 2 HTR 680		GO HDS : 400 VGO/DAO HDS: 1,190	
LHSV	Greater than 1.0		Greater than 1.0		VGO/DAO HDS: 0.15	
Commercial installation	Yes		Yes		Similar to atmospheric residue HDS unit of high severity	
Combustion of by-product						
Method	Pulverized coke combustion		Pulverized pitch combustion		Internal mixing steam atomizing	
Commercial installation	Many coke boilers in operation		Similar to pulverized coal combustion boilers		Straight run asphalt boilers are in operation. However, no experience of hard asphalt boiler	
Economics						
Capital requirements (10^6 US$)	1,073		1,097		1,188	
Return on equity, % (Income tax 50% case)	25.0		22.9		23.1	

(b) In the Eureka and M-DS processes, there are sections as shown below, where highly viscous liquids are handled at high temperatures.

- the pitch transfer system from the reactor bottom to a pitch flaker through a pitch stabilizer in the Eureka process.
- the asphalt rundown line from asphalt stripper and the liquid asphalt transfer system to boiler, as described in (3) "Combustion of by-product" in the M-DS case.

(c) The M-DS case requires large hydrogen consumption because the hydrodesulfurization of high severity is adopted. And, as many as 10 reactors of 500 ton class each are used for one hydrodesulfurization process unit.

(d) The Fluid coker and Eureka processes have commercial experience. Although many similar solvent deasphalting processes are in operation, the new extractor developed by Maruzen Oil Co. Ltd. has no commercial experience. As for VGO/DAO hydrodesulfurization, the unit has no experience of operation that gives a high rate of DAO.

(3) Combustion of by-product

Heavy residuals are inevitably produced in the course of the upgrading of heavy oil. Therefore, it is important to review the methods of converting the heavy residuals to energy by burning them on an industrial scale.

As a pulverized coke or coal combustion method is in commercial operation, the Fluid coker and Eureka cases which use the pulverized fuel combusition leave no problem.

In the M-DS case, it is proposed to use an entrainment type of combustion with high temperature steam. However in consideration of the fact that asphalt is handled over a temperature range in which low and medium pressure steams can not be used for heating purposes, operation may not be easy especially at the time of start-up and shut-down of the process.

2.5.2 Comparison from an Economic Viewpoint

Capital requirement of the M-DS case is bigger than those of other cases, because the hydrodesulfurization of high severity has to be adopted.

However, a high price of improved crude oil can be set for the M-DS case due to better gravity and sulfur levels. On the whole, there will be less difference in ROE (return on equity) among three cases.

CHAPTER 3 STUDY BASES

This chapter describes the study bases that have been set at the meeting with the Venezuelan authorities.

3.1 CRUDE OIL

The crude oil supplied to the Orinoco heavy crude oil upgrading refinery has been set as follows:

3.1.1 Raw Crude Oil

50/50% Cogollar IX and Cerro Negro Crude Oil

3.1.2 Crude Oil

(1) The crude oil supplied to the upgrading refinery is a mixture of the raw crude oil and the diluent which is required to pump up the raw crude oil.

(2) Distillate of 380–510°F, produced in the upgrading refinery, is used as the diluent.

(3) The ratio of diluent to raw crude oil is 0.3 on a volume basis.

3.1.3 Characteristics of Raw Crude Oil

Analysis data on the raw crude oil are based on "Crude assay of 50/50% COGOLLAR IX and CERRO NEGRO (Report No. LV. 5C–PC. 79)", included in Attachment 3 of VOLUME II. As to the data which are not described in the above assay, "TABLA 1 CARACTERIZACION DE LOS RESIDUOS (700°F+) Y DE SUS CRUDOS DE ORIGIN", included in ANNEX-C of Attachment 5, are to be referred.

3.1.4 Supply of Crude Oil

The crude oil produced at the field is given treatment such as water separation and desalting at the main station, and is supplied to the upgrading refinery battery.

Study on the main station is excluded from this study.

3.2. UPGRADING REFINERY

3.2.1 Location

The refinery will be located in the Cerro Negro oil field, south of Monagas.

3.2.2 Refinery Capacity

(1) Production capacity of the improved crude oil is 125,000 BPSD.

Feed rate of the upgrading refinery varies dependent on the process schemes.

(2) Stream days of the upgrading refinery are 330 days/year. The number of process trains and the intermediate tanks are to be studied in order to permit operation of minimum half capacity of process units during shutdown maintenance.

3.2.3 Configuration

Three cases utilizing the following processes, respectively, are to be studied.

(1) Fluid Coker Process

(2) Eureka Process

(3) M-DS Process

3.2.4 Utilities

(1) Self-supporting type to be studied.

(2) Required quantities of water for cooling and industrial uses are available from the river and will be supplied to the battery limits of the upgrading refinery.

3.2.5 Storage Facilities

(1) Crude oil tankage is set to have a capacity of 30 days storage.

(2) Improved crude oil tankage is set to have a capacity of 7 days storage.
The improved crude oil is pipelined to the shipping port.
Study of the storage capacity at the shipping port is excluded from this study.

(3) Capacities of intermediate tankage and by-product storage facilities are to be set by taking into account the operation of process units and boiler.

(4) Sulfur products storage facilities are set to have a capacity of one week's supply.

3.2.6 Desulfurization

Vacuum residue is mainly used as process fuel, and its flue gas is not desulfurized. Sulfur of 90% in the flue gas of boilers is recovered as gypsum or others.

3.2.7 Code and Standards

Internationally acceptable ones will be used.

3.3 IMPROVED CRUDE OIL

Improved crude oil (or synthetic crude oil) shall not include residual oil of the raw crude oil. Required properties of the improved crude oil are as follows:

Gravity: 25 – 28°API

Sulfur: 1 wt.% max.

Target yields of the improved crude oil and target key qualities of components are as follows:

3.3.1 Yield

Component \ Case	Thermal Cracking	Solvent deasphalting
C4/375°F	10–25 Vol.%	10–25 Vol.%
375/650°F	25 min.	25 min.
650/1,000°F	50 max.	40 max.
1,000°F+	0	25 max.

3.3.2 Qualities of Components

Components Qualities \ Case	Thermal Cracking	Solvent deasphalting
C4/375°F		
S, wt.%	0.05 max.	0.05 max.
N_2, ppm	2 max.	2 max.
375/650°F		
S, wt.%	0.2 max.	0.2 max.
Cetane No	35 min.	40 min.
650/1,000°F		
S, wt.%	0.7 max.	0.5 max.
N_2, wt.%	0.25 max.	0.10 max.
CCR, wt.%	1.0 max.	0.7 max.
Aniline Pt.	to be estimated.	
1,000°F+		
S, wt.%	N/A	1.25 max.

3.4 BY-PRODUCT FUEL

Residual oil and heavy liquid or solid product which shall not be included in the improved crude oil are used as boiler fuel.

3.4.1 Use

The by-products are utilized as fuel for the generation of electric power and steam for the upgrading refinery. The electric power is also supplied to the crude oil production field.

3.4.2 Electric Power for the Raw Crude Oil Production

150 MW is required for the raw crude oil production of 170,000 BPCD.

3.5 BY-PRODUCT SULFUR

Sulfur in the sour gas of hydrodesulfurization units is recovered as elemental sulfur.

Sulfur content of 90% in the flue gas of boilers is recovered as gypsum or others.

3.6 USE OF NATURAL GAS

Refinery off-gas and LPG are used as a feed for the hydrogen generation plant.

However, if the quantities of off-gas and LPG are less than the required quantity for the hydrogen plant, natural gas can be used.

3.6.1 Properties

C_1	93.1 mol. %
C_2	1.9 "
CO_2	3.7 "
C_3	1.3 "
Total	100.0 mol. %
H_2S	60 ppm
Mercaptan & COS	10 ppm

3.6.2 Supply Conditions

Pressure : 500 psig

Temperature : ambient

Required quantity is available.

3.7 COST ESTIMATION

Investment and operating costs are estimated on a Venezuelan site, 1980 basis.

3.7.1 Capital Requirements

(1) Import tax and duties of equipment : not included

(2) Fund : all equity

(3) Feed crude oil inventory : 15 days

(4) Improved crude oil inventory : 3.5 days

(5) Land cost : no value

3.7.2 Operating Cost

(1) Salaries including all allowances for operating personnel: US$22/MH

(2) Refinery organization

The upgrading refinery is organized by three departments: operation, maintenance and technical.

Other departments are organized outside of the upgrading refinery.

(3) For the scheduled maintenance, permanent maintenance personnel are to be supplemented by contracted maintenance personnel from outside of the upgrading refinery.

3.8 ECONOMIC ANALYSIS

An economic analysis is made by calculating the internal rate of return (IRR) based on assumed raw crude oil cost and improved crude oil price. The costs and prices are those of 1980 and escalation is not considered.

(1) Raw crude oil cost: US$10/BBL

(2) Improved crude oil price:

The price of Tia Juana Medium crude oil having the gravity very similar to the improved crude oil is set to be US$23.86/BBL and the following gravity differences and sulfur differences on a 1978 basis are used.

Gravity difference : US$0.08/API

Sulfur difference : for each 0.1% S in the 650°F fraction,

Range	Value
% S less than 0.5	US$0.25/BBL
% S between 0.5 and 1.0	US$0.15/BBL
% S between 1.0 and 1.5	US$0.08/BBL
% S between 1.5 and 2.5	US$0.04/BBL
% S greater tahn 2.5	US$0.02/BBL

Calculated values are then escalated to a 1980 basis at a rate of seven (7) percent per year.

(3) Sub-material cost and by-product cost: No value

(4) Electric power price: US$0.023/KWH

(5) Operating rate of the refinery:

1988 – 330 days/year x 50%

1989 – 330 days/year x 100%

(6) Others

Depreciation: 16.6 years, straight line method

Inventory, chemicals : 2 months

Income Tax : 50%

In addition to calculating the IRR based on the above conditions, sensitivity analyses are conducted for the following in order to review the effects of changes in the major factors which are taken as bases for calculating the IRR.

Change item	Base	Alternatives
Crude oil cost	US$10/BBL	± 50%
Construction cost	Base	± 20%
Income tax	50%	67%

Attachment to Study Bases

Estimation Bases of Process Data

The Process data in this study are estimated by the following bases.

	Fluid Coker	Eureka	M–DS
Main Upgrading Process	Fluid Coker	Eureka	M–DS
Yield	2 & 3	1 & 2	1 & 3
Properties	2 & 3	1 & 2	1 & 3
Operating Condition	2 & 3	1 & 2	1 & 3
HTR & HDS Process	HTR	No. 1 HTR	GO HDS
Yield	2 & 3	2	4
Properties	2 & 3	2	4
Operating Condition	2 & 3	2	4
		No. 2 HTR	VGO/DAO HDS
Yield		2	3 & 4
Properties		2	3 & 4
Operating Condition		2	3 & 4

Bases Number

1. Adjusted to Cogollar IX/Cerro Negro Crude bases by using the test data of the sample crude oil.
2. Estimated data based on the commercial operation of similar feedstock.
3. Estimated data based on the test of similar feedstock.
4. Estimated data based on various published information.
5. Data obtained from the licensers of HTR & HDS processes.
6. Others.

CHAPTER 4 PLANNING CONSIDERATION

This chapter describes the basic considerations which supplement the study bases defined in Chapter 3. The following are the contents of this chapter.

- Scope of study
- Product and by-products
- Pollution prevention
- Supply of diluent, fuel gas and electric power to raw crude oil production field
- Solid storage and handling facilities

4.1 SCOPE OF STUDY

The study covers the facilities included in the upgrading refinery consisting of process units, by-product combustion boiler, utility and offsite facilities.

Scope of the study on the upgrading refinery is shown by the red line in Fig. 4.1.

Facilities outside of the red line are excluded from the study.

The following flows are received or supplied at the red line.

Main input flow

Mixed crude oil
Natural gas
Industrial water
Limestone

Main output flow

Improved crude oil
Diluent
Fuel gas & By-product residual
Sulfur
Electric power
Waste water
Gypsum

The process units, the utility and offsite facilities, and the by-product combustion boilers are described in Chapter 5, 6 and 7, respectively.

4.2 PRODUCT AND BY-PRODUCTS

Main product of the upgrading refinery is the improved crude oil, whereas, the following are by-products:

- Sulfur recovered from acid gas of hydrodesulfurization units
- Gypsum recovered from boiler flue gas
- Electric power

- Refinery off gas
- Residuals of upgrading processes

4.2.1 Improved Crude Oil

Target qualities of the improved crude oil are suggested by the Venezuelan side, which are shown in Chapter 3. It is considered that the improved crude oil has sufficient competitiveness in the international crude oil market.

4.2.2 Sulfur Recovery

(1) Recovered by-product

Since the sulfur content of the Orinoco heavy oil is higher compared with those of other crude oils, desulfurization is required for the production of the improved crude oil. Also, when the by-product, in which sulfur in the crude oil is accumulated, is used as boiler fuel, SO_2 content in the boiler flue gas is very high.

The best removal method of sulfur is to produce sulfur by-product which can be used effectively. The second best plan is to produce sulfur by-product which causes no pollution problem. In addition to those, the sulfur removal method must be inexpensive. Sulfur compounds in the flue gas can be removed by water washing; however it results in water contamination which does not mean pollution control.

Therefore, kind of sulfur by-products should be studied.

In the case of upgrading the Orinoco heavy oil large quantity of sulfur by-product is produced; therefore, the sulfur recovery method should be studied in consideration of the uses of and demands for sulfur by-products.

The largest use of sulfur compounds is sulfuric acid, but it is a by-product of smelting of non-ferrous metals. Therefore, large demand is not expected for the recovery of sulfuric acid in the case of the upgrading of the Orinoco heavy oil.

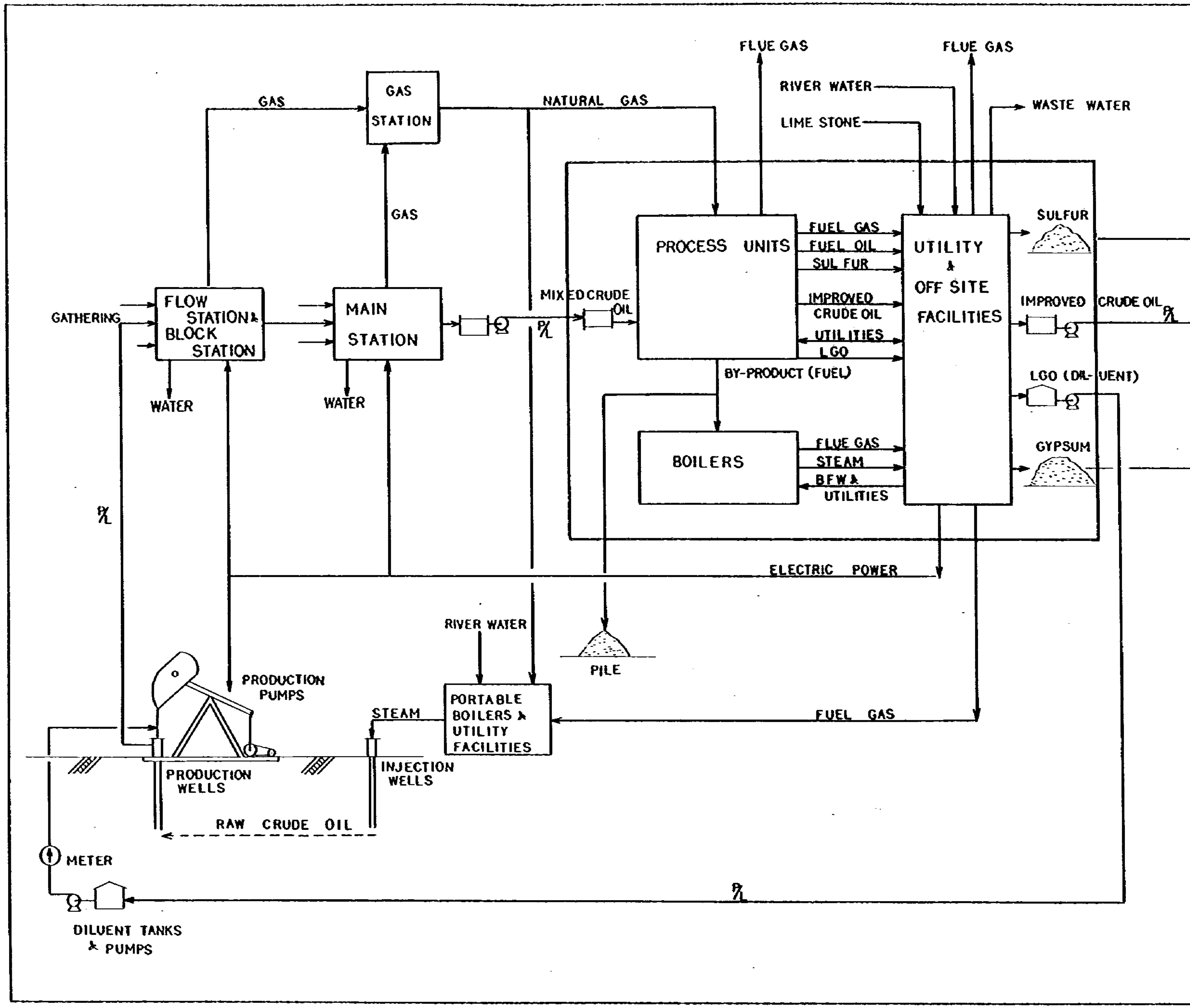
Demand for carbon disulfide, paper, pulp, sodium sulfite and sodium sulfate is not so large, while demand for gypsum comes next to that for sulfuric acid, and is expected to grow in future. Gypsum is used as such noninflammable construction materials as gypsum board and gypsum plaster, and also is mixed with cement.

In this study, the sulfur recoveries are studied based on the following conditions.

- Elemental sulfur, which is a material for the sulfuric acid production, is recovered from the off gas of hydrodesulfurization processes.
- Gypsum, which is expected to have large demand and is a non-pollution material, is recovered from the boiler flue gas.

(2) Form of elemental sulfur

Forms of elemental sulfur are liquid (molten sulfur) and solid (pelletized sulfur). Major characteristics of sulfur are shown in the table below.



: BATTERY LIMIT OF FEASIBILITY STUDY

P/L : PIPE LINE

BFW : BOILER FEED WATER

LGO : LIGHT GAS OIL

ORINOCO HEAVY OIL UPGRADING PROJECT

FRAME OF PROJECT

FIG. 4.1

As is shown in the table, the fluidity of sulfur is high in the temperature range of 120°C–155°C; however, above this temperature, the fluidity decreases sharply and reaches at its peak at 190°C. The handling temperature of molten sulfur is generally about 140°C in consideration of its viscosity characteristics as well as safety control for ignition by impurities such as hydrogen sulfide.

Characteristics of Sulfur

Viscosity:

Temp. (°C)	Vis. (cp)	Temp. (°C)	Vis. (cp)
120	10.1	180	29,600
130	9.26	190	33,200
140	8.40	200	31,700
150	7.76	220	20,300
158	7.90	240	12,200
160	30.2	260	6,150
165	5,380	300	2,020
170	15,900	340	530

Specific Gravity:

Temp. (°C)	Sp. Gr.
121.1	1.804
154.4	1.775
178.3	1.767
210	1.751
357	1.658

Melting Point: 118.9°C

Freezing Point: 114.5°C

Vapor Pressure:

Temp. (°C)	Press. (mmHg)	
114.5	0.0285	Solid
123.8	0.0535	
141.0	0.131	
177.0	0.625	
211.3	3.14	
352.5	133.0	

It is planned that the sulfur is shipped in pelletized form, taking into consideration such liquid handling problems as storage at high temperature, transportation by pipeline and shipment by tanker.

4.2.3 Combustion of By-Products

Utilization of the by-products is described in Chapter 7. In the study, the by-products are utilized as boiler fuel and a surplus by-product which exceeds the boiler fuel requirements is stock-piled.

4.2.4 Diluent

As the Orinoco heavy crude oil has high viscosity, diluent is required to reduce the viscosity for pumping of the heavy oil.

The diluent is set to be light gas oil (Boiling range is 380–510°F) recovered from atmospheric crude distillation units in the upgrading refinery and the ratio of diluent to raw crude oil is set at 30 percent on a volume basis.

Estimated viscosities of the raw crude oil, the diluent and the mixed crude oil are as follows:

	Viscosity (cst.)		
Temp. (°F)	Raw Crude Oil	Diluent	Mixed Crude Oil
100	95,000	2.25	1,300
130	10,517	1.70	350
140	5,944	1.58	245
180	831	1.18	71
210	270	0.99	33

The viscosity of mixed crude oil is the same level as those of long residue of the Middle East crude.

Crude	Kuwait	Arabian Light		Arabian Heavy	Iranian Light	
Fraction	650°F+	650°F+	725°F+	650°F+	650°F+	725°F+
Viscosity (cst.)						
@ 100°F	980	280	700	5,500	490	1,450
@ 122°F	420	140	310	2,000	220	750
@ 210°F	42	19.5	34	118	26	46.5

4.3 POLLUTION PREVENTION

As, at present, there are no restrictions and regulations in Venezuela for pollutions control, the pollution prevention facilities are planned based on the following models:

4.3.1 Water Pollution Prevention

Waste water discharged to the adjacent river finally flows to the Paria gulf.

Therefore, regulations on water pollution should be set in consideration of river and sea water.

In this study, for the time being, the facilities of foul water stripper and oil separator as the first treatment are planned to be installed in the refinery. And the ballast water from tankers is not included because the receiving and loading facilities for tanker are excluded from the study.

4.3.2 Air Pollution Prevention

Harmful materials in the waste gas discharged from the upgrading refinery affect plants, vegetables and human being.

Agricultural products and trees are scarce in the Orinoco area, and distance and direction of residential facilities from the upgrading refinery are not clear. Therefore, regulation on the waste gas can not be fixed. There is a possibility that requirements can be satisfied by only high stack.

In this study, the facilities for air pollution prevention are planned based on the following models:

(1) Hydrogen sulfide included in the gas produced by such processes as hydrotreater and hydrodesulfurization is recovered as elemental sulfur.
However, sulfur recovery from tail gas of a sulfur recovery unit is not planned.

(2) Sulfur dioxide included in the flue gas of process furnaces using vacuum residue as main fuel is not recovered.

(3) Sulfur compounds content of 90% in the flue gas of boilers using the by-products as fuel is recovered by the flue gas desulfurization units.
Facilities for nitrogen oxide removal are not planned, for the time being.

4.4 SUPPLY OF DILUENT, ELECTRIC POWER AND FUEL GAS TO RAW CRUDE OIL PRODUCTION FIELD

Diluent, electric power and fuel gas for the production of injection steam required for the raw crude oil production are to be supplied from the upgrading refinery.

As is described in Chapter 3, the study bases are:

- stream days of the process units are 330 days/year, and

- minimum half capacity of process unit is to be operated during the shutdown period (365 – 330 = 35 days).

Therefore, the following considerations are made to keep stable supply of diluent, electric power and fuel gas to the crude oil production field:

4.4.1 Diluent

Diluent storage tanks are required to supply diluent continuously for the production of raw crude oil. The tanks will be installed in the upgrading refinery and/or in the crude oil production field. In the study, for the time being, the planned capacity of the diluent tanks in the refinery is 7 days' production of the diluent.

4.4.2 Electric Power

As by-products (coke and pitch) are used as fuels for the generation of steam, facilities relating to the generation and supply of electric power are planned by taking into consideration the characteristics of by-products and the combustion methods which are described in Chapter 7.

The following are basic items for the facility plans on the bases of respective combustion methods. The results of the study are described in Chapters 5, 6 and 7.

(1) Fluid coker case

Combustion method:

Pulverized fuel combustion with supporting fuel

Study items:

- spare boiler and generator
- coke storage facility
- number of trains of the Fluid coker process
- supply of supporting fuel

(2) Eureka case

Combustion method:

Pulverized fuel combustion

Study items:

- spare boiler and generator
- the Eureka pitch storage facility
- number of trains of the Eureka process

(3) M-DS case

Combustion method:

The M-DS pitch, which is directly sent from the process to the boilers with high temperature, is atomized by high temperature steam.

Study items:

- number of trains of the M-DS process
- use of vacuum residue as an alternative fuel and vacuum residue storage facility

4.4.3 Fuel Gas

Field portable boilers for the generation of injection steam are installed separately in the raw crude oil production field. Main fuel for the boilers is natural gas. However, supply of off gas generated in the upgrading refinery to the field will contribute to the refinery economy.

In order to supply the off gas to the field, fuel balance in the upgrading refinery is planned on the following bases:

(1) Vacuum residue is used as main process fuels.

(2) Use of the off gas as process fuels is limited to hydrogen plant and others which require gas or light liquid fuels.

4.5 SOLIDS STORAGE AND HANDLING FACILITIES

With the upgrading of Orinoco heavy oil, it becomes necessary to store and handle the solid form of raw materials and products. Their handling amounts are much more than conceivable for ordinary refineries, and can be even enormous, depending upon facility-operating and product-shipping conditions. The following points call for the solids handling at the refinery under study:

- All of the by-products (heavy residuals) are in the solid form at normal temperature, when heavy crude oil is processed by the upgrading process.
- Because crude oil to be used contains a high level of sulfur, the desulfurizer by-produces sulfur during the course of production of improved crude oil having a low sulfur content.
- When a by-product with a high sulfur content is used as fuel, the desulfurization unit of boiler flue gas also by-produces solids (such as gypsum and sulfur).
- Limestone is used as a raw material if gypsum is byproduced by the flue gas desulfurization.

Countermeasures must be taken to cope with these points, now that they are the inevitable to be accepted by any upgrading refinery.

4.5.1 Volumes of Solids Storage

Actual solids storage volumes are determined on the basis of prerequisite conditions for each of the three cases now being studied. An illustrative model of the types and volumes of by-products is given as follows:

(1) By-products (heavy residuals)

By-products from the three process cases are:

Fluid coker case .... Fluid coke

Eureka case .... Eureka pitch

M-DS case .... M-DS asphalt

The above by-products other than fluid coke can be stored outdoors. For the convenience of simple calculation, here are placed the following assumptions: Either one of these by-products is produced at a rate of 4,500 T/SD; boiler can be supplied with fuel even at the time of periodical maintenance; apart from the supplies to the boiler, the surplus amount of the by-product is stored in the refinery in an amount corresponding to a month of production; and any further surplus in excess of a month of production is stock-piled elsewhere outside the refinery. Based on these assumptions, the amount stored within the refinery is calculated as:

4,500 T/DS x 30 days = 135,000 tons

With the specific gravity set at 1.0 for the convenience of simplicity, the above amount will form a pyramid with a square base of 200 m and a height of 10 m.

(2) Sulfur

If crude oil is hydrodesulfurized to remove sulfur, which is recovered in the form of elemental sulfur, such sulfur is solidified for the convenience of storage and transportation. If it is produced at a rate of 500 T/SD, a week of production to be stored within the refinery is calculated as:

500 T/SD x 7 days = 3,500 tons

With the specific gravity of sulfur set at 1.8, this amount gives a volume of about 2,000 m^3, which forms a pyramid with a square base of 35 m and a height of 5 m.

(3) Limestone and gypsum

Desulfurization of boiler flue gas depends on the boiler capacity. If it is assumed that boiler flue gas is desulfurized at a rate of 150 T/SD in terms of sulfur, 470 T/SD of limestone is used to give 800 T/SD of gypsum. If limestone is stored in an amount sufficient to operate the flue gas desulfurization process for 30 days, the limestome requirement is calculated as: 470 T/SD x 30 days = 14,100 tons. If a week of gypsum production is stored, the gypsum storage amount is obtained by: 800 T/SD x 7 days = 5,600 tons. Since both of limestone and gypsum are not stored outdoors, a warehouse or respective silos must be installed.

4.5.2 Facilities for Solids Receiving, Storage and Shipment

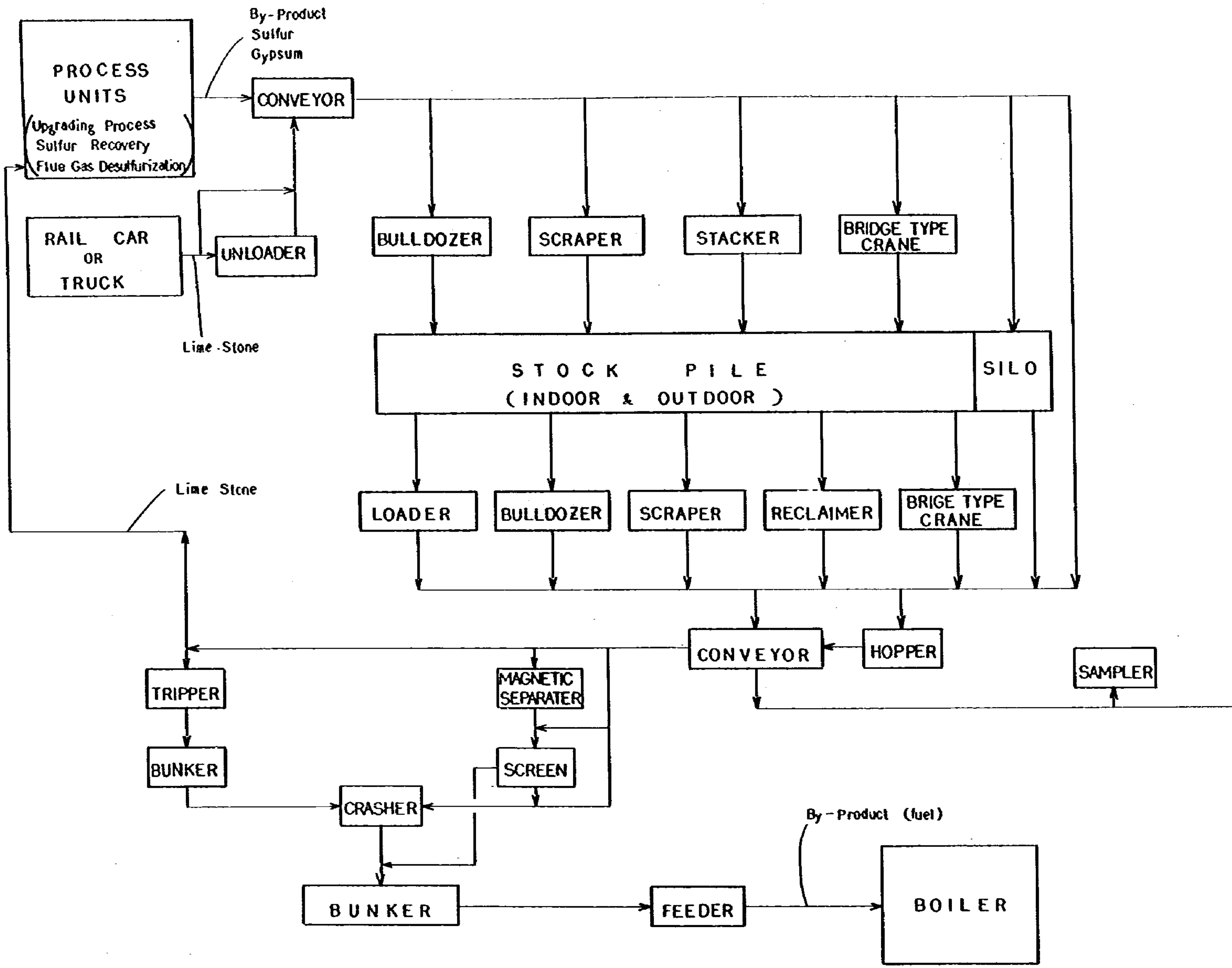
Optimum methods must be selected to receive, store and ship the solid materials, taking into consideration the volumes of handling and storage, methods of storage, the terrain and layout of outdoor stock-piling ground, and whether storage is indoor or outdoor. Fig. 4.2 shows a solids flow diagram among the facilities. If consideration is given to the afore-mentioned receiving of limestone from outside and transportation of sulfur and gypsum to the shipping port, it is premised that transportation occurs only in the daytime and that railroad is the best means of transportation in the light of large daily production and shipment. However, this study excludes the problem of selecting an optimum means of transportation, anticipating that the matter will be dealt with under a plan on infrastructures.

As stated above, the by-product of heavy residuals from the upgrading process is partly consumed as boiler fuel, and the surplus amounts keep piling up as long as the refinery is in operation. Obviously, all of these surplus amounts cannot be stored infinitely in the refinery.

It is thus assumed that, except for the amount required for boiler operation, all the surplus is stock-piled elsewhere outside the refinery. Outside storage and transportation to the storage site are also excluded from this study.

For reference, the following data indicate the sizes of gondola cars commonly used for railroad transportation of coke in Japan.

Class	TORA	TOKI
Load limit, MT	17–18	35
Capacity, m^3	37.8–44.5	66.7–67.7
Length, mm	2,450–8,650	12,700–12,900
Width, mm	2,450–2,585	2,500
Depth, mm	800–970	965



ORINOCO HEAVY OIL UPGRADING PROJECT

STORAGE AND HANDLING OF SOLID

FIG. 4.2

CHAPTER 5 DESCRIPTION OF PROCESS UNITS

This chapter describes outlines of process units to be used in the Orinoco heavy oil upgrading refinery being studied. These process units are examined in the following three cases in which they are used for upgrading:

Fluid Coker Process
Eureka Process
M-DS Process

Descriptions in this chapter are based on the data given in Chapters 1, 2 and 3 of Volume II, which are the results of study made by all the process- proposing companies under the same prerequisite conditions.

5.1 UPGRADING PROCESS UNITS

This section outlines the upgrading processes proposed by the three companies, i.e., Fluid coker process, Eureka process, and M-DS process.

For detailed information, refer to Sections 1.1, 2.1, and 3.1 of Volume II.

5.1.1 Fluid Coker Process

(1) Introduction

The fluid coker process is a continuous thermal cracking process developed by, and patented to, Exxon Research & Engineering Company (hereinafter referred to as ERE) for the thermal conversion of heavy oil of low quality to light hydrocarbons and coke. There is no other process of this type that has been commercialized or will be developed in the near future.

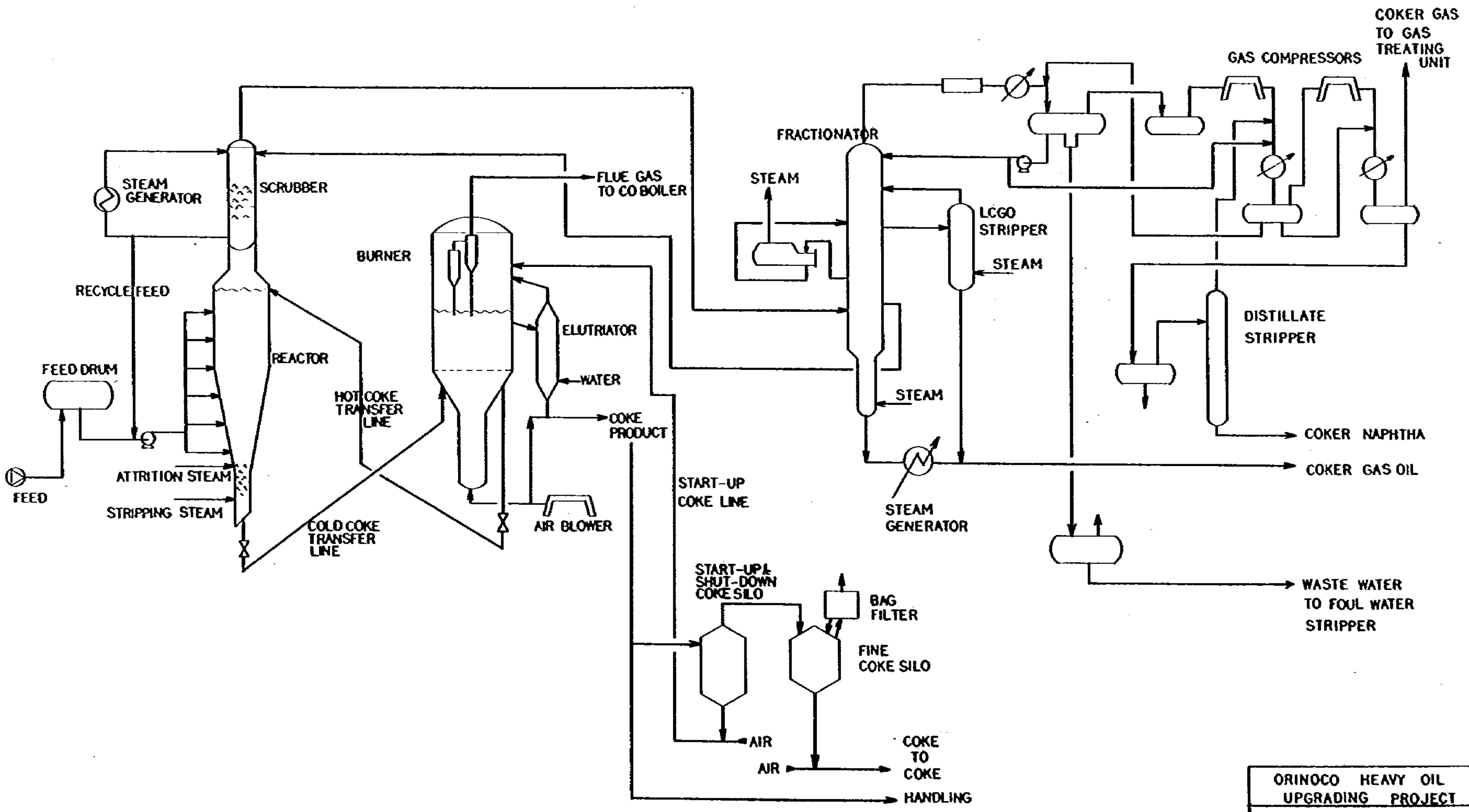
In this process there is used a fluidized-bed technology that has been developed as a fluidized catalyst bed cracking technology. However, instead of using a catalyst, the fluid coker process employs pulverized coke particles (50 to 2,000 microns), which are circulated between the reactor and the burner at a high temperature. Because oil is thermally cracked over the fluidized bed of coke particles circulated at a high temperature, the process does not need a furnace, and the plant based on this process can be run under severe operating conditions.

The first plant based on this process went into operation in 1954, and ever since more than 10 plants are in successful operation. Table 5.1 gives world experiences in the fluid coker and the flexicoker which will be described later.

ERE later developed the flexicoker process, wherein a gasifier is added to the fluid coker process to convert the coke by-produced in the fluid coker to fuel gas so as

Table 5.1 Experiences in Fluid Coker & Flexicoker

Company	Location	Design Capacity (B/SD)	Date Onstream	Remarks
FLUID COKER				
Humble Oil Co.	Billings, Mont.	3,800	Dec. '54	
Humble Oil Co.	Baltimore, Md.	10,000	Sept. '55	Entire refinery shut down
Petrofina Canada Limited	Pointe-aux-Trembles, Quebec	3,800	Aug. '56	
Marathon Oil Co.	Detroit, Mich.	4,000	Oct. '56	Plant formerly operated by Aurora Gasoline Co.
Signal Oil Co.	Bakersfield, Calif.	4,000	Apr. '57	Plant formerly operated by Bankline Oil Co.
Phillips Petroleum Co.	Associated. Calif.	42,000	June '57	Plant formerly operated by Tide-water Oil Co.
Getty Oil Co.	Delaware City, Del.	42,000	Aug. '57	(")
Gulf Oil Corp.	Purvis, Miss.	4,800	Dec. '57	Plant formerly operated Pontiac Eastern Corp.
Petroleos Mexicanos	Madero, Mex.	12,000	Feb. '68	
Imperial Oil Ltd.	Sarnia, Ontario	15,000	Apr. '68	
Humble Oil Co.	Benicia, Calif.	16,000	Apr. '69	
Syncrude Oil Co.	Mildred Lake, Alberta, Canada	73,000 73,000	July '78 '78	
FLEXICOKER				
Toa Oil Co.	Kawasaki Japan	21,300	Sep. '76	
Lagoven	Amuay Venezuela	52,000	----------	Onstream in 1982 under construction
Imperial Oil Co.	Coldlake Alberta	50,000	----------	Under constructin
Unannounced	Canada	One	----------	Under design
Unannounced	U.S.A.	Two	----------	Under design



ORINOCO HEAVY OIL UPGRADING PROJECT

PROCESS FLOW DIAGRAM OF FLUID COKER PROCESS

FIG. 5.1

to meet the demand for fuel of an area concerned. Toa Oil Co., Ltd. owns the first plant based on the flexicoker process, which is the only plant now in operation in the world. With its many knowhows, Toa investigated which of fluid coker and flexicoker is beneficial for the project now being studied, and proposed the fluid coker for the project. The fluid coker has the following features:

(a) A wide range of raw materials, from light to heavy ones, can be used if they are supplied by pumping.

(b) Attrition steam jets have been developed for the pulverization of seed coke particles, although ordinary methods can be used for this purpose.

(c) Process equipment is mainly composed of a reactor and a burner. The reactor allows coking and a part of coke is burnt in the burner to supply reaction heat necessary for the process.

(d) Coke is circulated between the reactor and the burner in such a way that density of the fluidized bed is adjusted by the differences in pressures and height between the reactor and the burner.

Aeration steam is ejected into the coke transporting line to aid the circulation.

(e) The reactor and the burner are operated at temperature ranges of 850–1,000°F and 1,000–1,200°F, respectively.

(f) The amount of coke used is determined by the amount of Conradson carbon and the reactor operating conditions.

(g) The higher the cut point temperatures of circulated oil, the lower the coke yield becomes.

(h) The raw material in an amount of 5–10% is usually burnt and used as the process heat source.

(i) Economy of the fluid coker process is improved by raising the gas oil production, but 1,050°F will be a reasonable limit.

(2) Description of process flow

Fig. 5.1 shows the process flow of the fluid coker process.

The feed of heavy oil pre-heated to a high temperature is sprayed on the circulated fluidized coke particles by steam injection. The feed deposited on the surfaces of coke particles is then thermally cracked to give light hydrocarbons and coke. The coke formed remains on the circulating coke and goes to the burner. The cracked light hydrocarbons are cooled by scrubber in the top part of the reactor, and at the same time, are gotten rid of entrained coke. The hydrocarbons are then passed into distillation section, where the hydrocarbons are separated into cracked gas and cracked oil.

Uncracked heavy oil in the reactor is condensed by the scrubber and recycled to the feed line for further cracking in the reactor.

Stripping steam is fed to the reactor through the bottom to remove hydrocarbons from the surfaces of coke and at the same time, to form a fluidized bed of the reactor. In addition, attrition steam is also fed to the reactor to secure successful coke circulation by crushing coke lumps and adjusting particle sizes of fluid coke.

The heat requirement of the reactor is met by circulating high-temperature coke coming from the burner. Heat is generated at the burner by taking in air and burning part of the circulated coke. Heat balance is maintained for the equipment by controlling the volume of air taken into the burner.

Net surplus coke not used for combustion as well as for circulation is withdrawn from elutriator as a by-product.

The burner generates a low-calorie gas of a high temperature, which is normally used for steam generation at CO boiler.

A series of fluid coker is filled with 2,500 tons of coke. Because the plant is started up, with a train at a time, only a silo with a 3,000 ton capacity will be sufficient for operation startup and shutdown.

(3) Process yields

Table 5.2 gives estimated yields of products from the fluid coker process.

Using vacuum residue as the raw material for the fluid coker, the process produces 62.48 vol. % (or 53.35 wt. %), based on the volume (or weight) of the feed, of cracked naphtha and gas oil, with 26.60 wt. % of coke being by-produced. The balance includes cracked gas and CO gas, both used as fuel gases.

5.1.2 Eureka Process

(1) Introduction

Eureka process was developed as a thermal cracking process producing light cracked oil easily capable of desulfurizing residue oil. This cracking process by-produces pitch, which is used for various purposes. The Eureka process was developed by, and patented to, Kureha Chemical Industry Co., Ltd. Commercial plants based on this process are now in successful operation. To commercialize the newly developed process, there was established Eureka Industry Co., Ltd. in 1972. The company constructed a plant at a site facing Tokyo Bay in Chiba Prefecture, Japan. The plant, the first of its kind, processes about 20,000 BPSD of vacuum residue, and produces about 14,400 BPSD of cracked oil which is desulfurized to give fuel oil, and also by-produces about 300,000 tons/y. of pitch to be used as a binding agent for metallurgical coke.

Another plant based on this process is now under construction in the People's Republic of China.

Table 5.2 Yield of Fluid Coker Process

Feed	vol.%	°API	wt.%
Orinoco			
Vacuum Residue (995°F+)	100.0	1.8	100.0
Products			
Gas (C_4–)	(18.8 Nm^3/B)	–	(13.35)
Naphtha (C_5/360°F)	15.67	60	10.90
Gas Oil (360/950°F)	46.81	15.5	42.45
[Coke (Gross)]	–	–	[33.30]
Coke (Net)	–	–	26.60
CO Gas	–	–	6.70 Equivalent

Properties

C_3^- Gas		C_4 Gas		Naphtha	Gas Oil
H_2	16.7 mol.%	$C_4^=$	60 mol.%	0.84 wt.%S	3.38 wt.%S
C_1	44.8 "	C_4	40 "	Br. No 130	29 cst @ 110°F
$C_2^=$	7.7 "		100 "		18 cst @ 130°F
C_2	15.6 "				0.6 wt.% N
$C_3^=$	8.1 "	MW=56.8			4.0 wt.% CCR
C_3	7.1 "	LHV=28,100 Kcal/Nm^3			1.96 ppm V
	100.0 "				1.05 ppm Ni
MW=20.9					0.12 ppm Fe
LHV=11,050 Kcal/Nm^3					Br. No 43

CO Gas		Coke	Mesh Size Normal	% ON	Fine Coke	% ON
CO_2	12.0 mol.%	5.79 wt.%S				
CO	6.0 "	2,460 ppm V				
N_2	56.5 "	610 ppm Ni	20 (841μ)	5	1,000μ	3.2
H_2S	0.5 "	70 ppm Fe	50 (297μ)	15	590μ	0.2
H_2O	25.0 "	Bulk density	60 (250μ)	25	297μ	0.4
		56 Lb/CF	80 (177μ)	55	250μ	1.3
	100.0 "	(0.897 g/cm^3)	100 (149μ)	65	177μ	5.5
MW=27.4			140 (105μ)	75	149μ	15.6
(Dry Base=30.6)			200 (74μ)	95	125μ	24.9
LHV=242 Kcal/Nm^3					105μ	20.7
Dust 0.213 T/H					88μ	12.5
					74μ	5.0
					63μ	6.7
					53μ	2.9
					53μ⁻	1.1

Most of the petroleum residues are intricate mixtures of hydrocarbons of unknown structures which are difficult to describe in the chemistry of thermal cracking processes. Fundamentally, though, their chemical reactions are considered to include the following two reactions:

- Both paraffinic and naphthenic hydrocarbons of large molecular weights are thermally cracked to give paraffinic oils of low molecular weights containing small amounts of gaseous hydrocarbons.
- Radical condensation reaction between dealkylated asphaltenes forms higher molecular weight aromatic pitch.

These reactions are continuously carried out in a liquid phase.

The features of Eureka process are as follows:

(a) A high yield of cracked oil is obtained.

(b) The reactor bottom material (pitch) can be handled in a fluid state.

(c) The produced pitch is a highly stable and homogeneous material.

(d) The cracked gas yield is low at only 4 to 5%.

(2) Description of process flow

Fig. 5.2 shows a process flow diagram for the Eureka process.

The feed oil is preheated to 350°C by a preheater if it has a low temperature at the time of feeding, and is fed to the main distillation tower through the bottom and mixed with a circulated heavy fraction of the cracked oil. The ratio of circulated oil to newly fed residual oil is normally in the range of 0.2 to 0.3.

When the residual oil is fed to the main fractionator through the bottom, it is pre-heated to about 500°C at a charge heater and fed to one of the reactors, as directed by an automatically working switch valve. Reaction proceeds semi-batch-wise at a preset cycle.

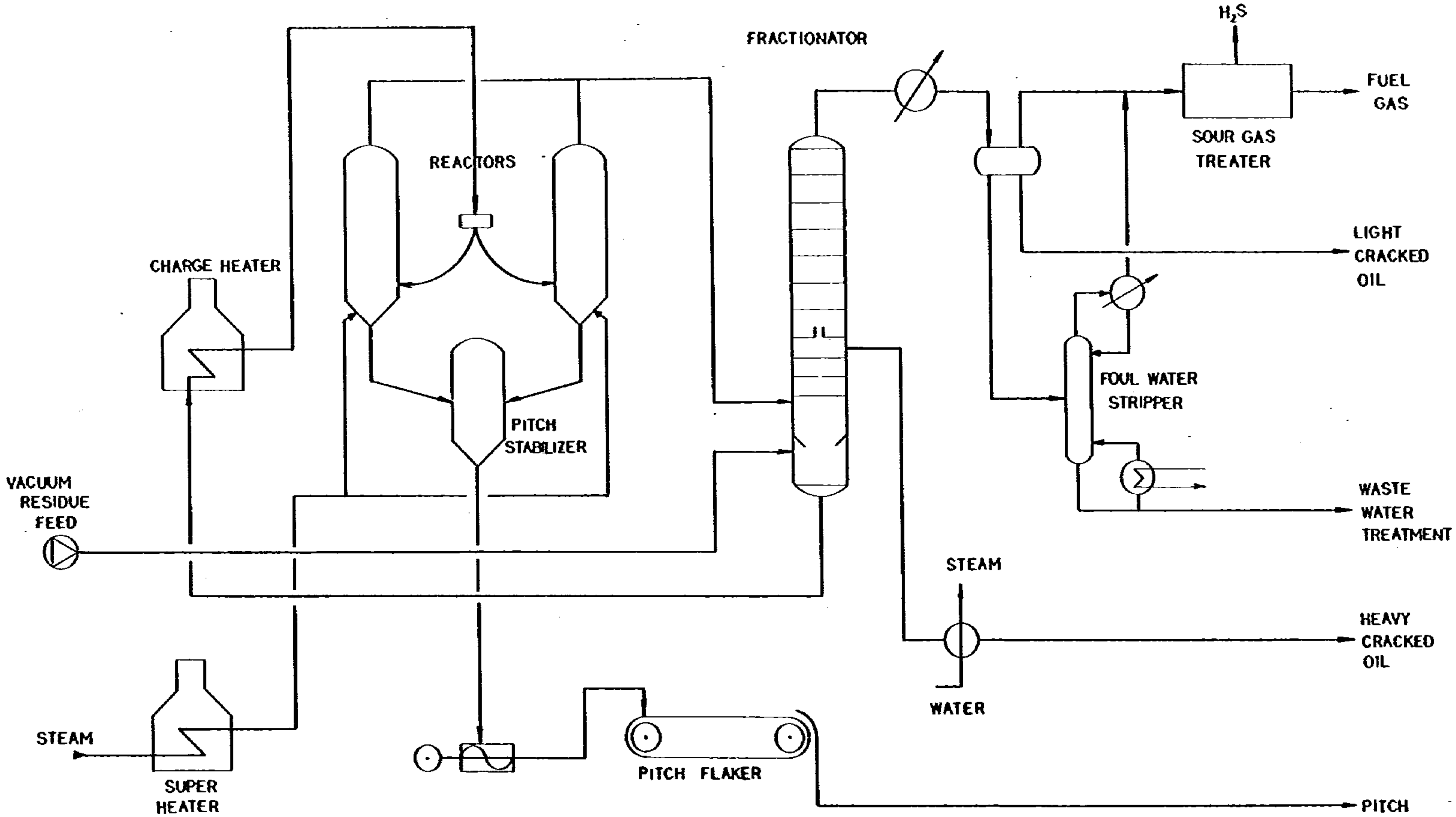
A typical 4-hour cycle involves:

– Raw material feeding	About 2 hr.
– Soaking	About 1 hr.
– Quenching, blowdown and stand-by	About 1 hr.

Thus, the feeding operation can be switched between the two reactors at a time interval of 2 hr.

If the process is operated on a large scale, either the number of process trains is increased, or the number of reactors is increased so that several sets of the above-described cycle can be performed. In such a case it will be convenient to computer-control the switching operation.

Steam, superheated to 600–700°C, is fed to these reactors to provide them with reaction heat and to strip thus-cracked oils. A reaction temperature of 400 to 450°C is used.



ORINOCO HEAVY OIL UPGRADING PROJECT

PROCESS FLOW DIAGRAM OF EUREKA PROCESS

FIG. 5.2

With the progress of reaction, cracked oil is stripped off, and the residue undergoes condensation and polymerization to form viscous pitch. When the pitch in the reactor reaches a certain constant flow point, the reaction is terminated by spraying water directly into the reactor to cool the temperature inside the reactor down to 300–350°C.

Pitch obtained is dropped by gravity into pitch stabilizer below the reactor, where a small amount of steam is fed to get rid of volatile components, and pitch is then passed into pitch flaker.

The lines and pump used for pitch transfer are designed to keep pitch flow-ability by means of a heat-stable heating medium.

Cracked oil and gas are sent to the main fractionator, where they are separated from entrained pitch and heavy fraction and the oil is further fractionated into light cracked oil and heavy cracked oil.

Cracked gas is passed to a treatment unit for the use as a fuel gas.

A large amount of steam is generated through heat recovery from cracked heavy oil.

(3) Process yields

Estimated yields from the Eureka process are given in Table 5.3, below.

Using Orinoco vacuum residue as the feed, the Eureka process gives 67.85 vol. % (or 59.0 wt. %), based on the volume (or weight) of the feed, of cracked oils and 37.0 wt. % of by-produced pitch. Cracked gas is given in an amount of 4.0 wt. %, based on the feed, which percentage is less than given by the fluid coker process. This amount is totally used as fuel gas.

5.1.3 M-DS process

(1) Introduction

Solvent deasphalting process is one of the fundamental refining technologies, which has been used for more than 50 years in the petroleum industry. It was in early 1970s that Maruzen Oil has commenced on the research and development of M-DS process. In 1972 the company constructed a large pilot plant on the basis of about 2 years' fundamental studies.

Since then, the company has continued test runs with this pilot plant for about 4 years. In 1976, basic design has been completed for the construction of a full-scale plant using 33,000 BPSD of residues from Middle East crude oils. Patents for the M-DS process is under application.

Any type of feed can be used for the process, as long as the feed is a heavy oil containing a residue component. Solvent deasphalting gives deasphalted oil of improved quality. Since, however, this process is most effectively used as a pre-treatment process, suitable feed may be atmospheric or vacuum residues having

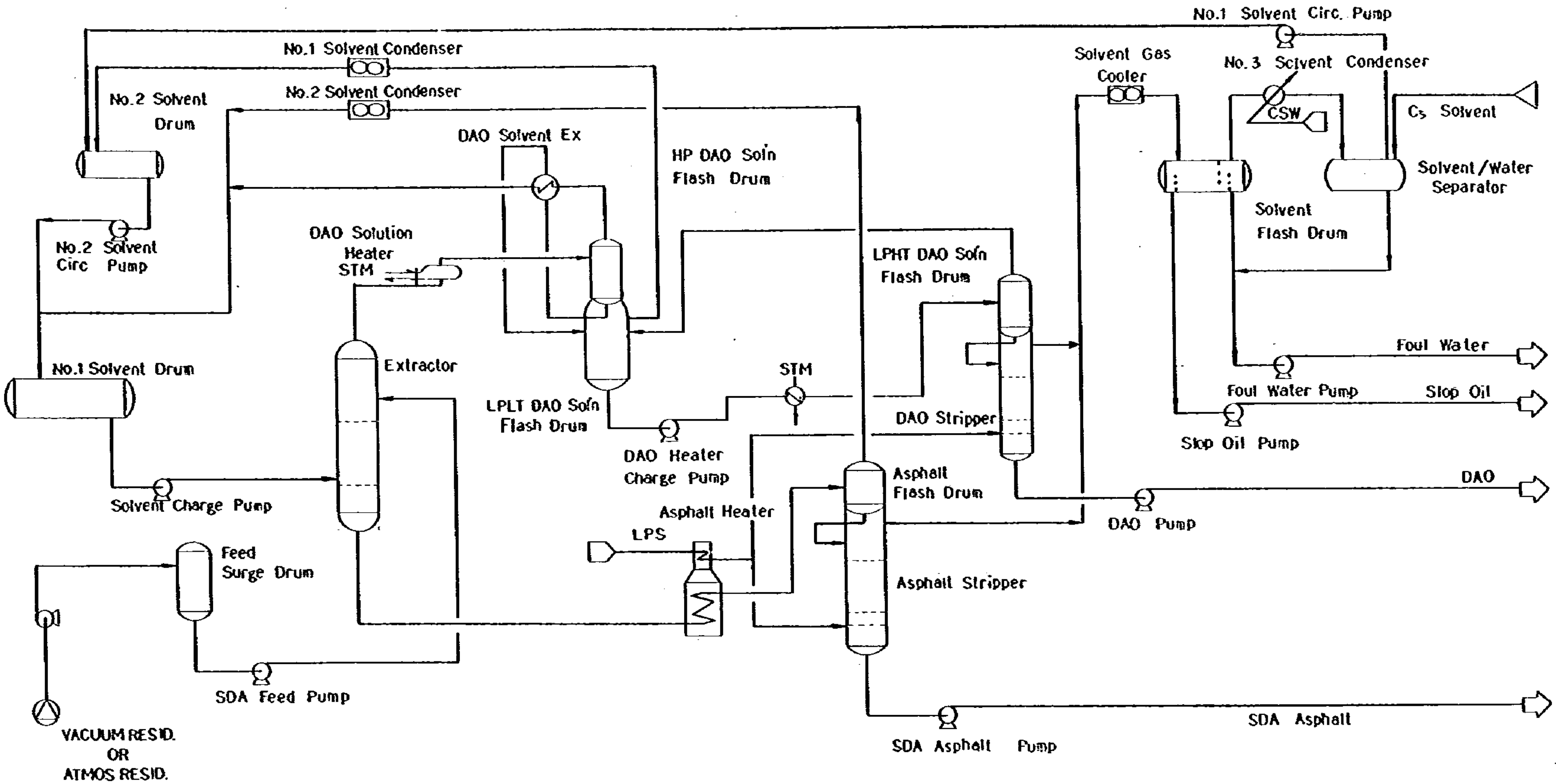
Table 5.3 Yield of Eureka Process

Feed	vol. %	°API	wt. %
Orinoco			
Vacuum Residue (995°F)	100.0	1.8	100.0
Products			
Gas		–	4.0
Cracked light oil	11.44	52.3	8.3
Cracked heavy oil	56.41	17.0	50.7
Pitch	–	–	37.0

Properties

Cracked Gas		
H_2	3.54	vol. %
CH_4	35.42	"
CO	1.73	"
CO_2	1.39	"
C_2H_4	2.07	"
C_2H_6	15.16	"
C_3H_8	8.75	"
C_3H_6	4.62	"
C_4H_{10}	4.87	"
C_4H_8	4.82	"
1.3-C_4H_6	0.01	"
H_2S	17.59	"
RSH	0.3	"

		CLO	CHO
Normal Cut pt.	°F	C_5-482	482–1,000
API	–	50.6	16.8
Sulfur	wt. %	0.6	3.6
Nitrogen	"	–	0.3
Vanadium	ppm	–	<0.1
Nickel	"	–	<0.1
Bromine No.	–	84.9	42.3
Diene Value	–	4.9	4.5
Total Acidity	–	<0.1	1.46
ASTM Dist.			
IBP vol. %, °F		109	426
10 " , "		205	576
50 " , "		324	788
90 " , "		442	928
97 " , "		486	995



ORINOCO HEAVY OIL UPGRADING PROJECT

PROCESS FLOW DIAGRAM OF M-DS PROCESS

FIG. 5.3

high contents of heavy metals and asphalten.

Solvent deasphalting is in principle regarded as a pretreatment technology. The process involves deasphalting of the feed to reform it. It is widely used as a feed adjusting method in such processes as lubricant refining, hydrocracking, and catalytic cracking. Asphalt contains most of highly condensed hydrocarbons and metallic compounds detrimental to the desulfurizing catalysts. Vanadium and nickel can be easily removed by solvent deasphalting.

Industrially used solvents include light paraffinic hydrocarbons, such as propane, butane and pentane. The M-DS process has the following features over other solvent deasphalting processes:

(a) The yield of deasphalted oil is higher.

Development of a high-performance extractor has enabled deasphalted oil to be obtained at a high yield by the use of a low solvent ratio, a low deasphalting temperature, and relatively high molecular weight solvents.

(b) Costs of plant construction and operation are less expensive. Development of a high-performance extractor and plant optimization could have reduced these costs.

(2) Description of process flow

Fig. 5.3 shows the process flow of the M-DS process. The feed, i.e., either atmospheric residue or vacuum residue, is mixed with a solvent and fed to the extractor. A part of solvent is separately fed to the extractor through the bottom, to fully extract the oil remaining in the asphalt fraction. The mixture of feed and solvent is divided into an extract phase and a raffinate phase in the specially designed extractor. The extract phase, consisting of oil and solvent, is heated by an extract (DAO Solution) heater, and sent to the extract (DAO Solution) flash drum, where solvent is separated from oil and, after cooling, recycled to the solvent drum No. 2. The deasphalted oil is heated by the DAO heater and passed into the DAO stripper, where a trace amount of solvent remaining in the oil is thoroughly stripped off, and DAO free of solvent is transferred to the desulfurization section.

In the meantime, the raffinate phase, consisting of solvent and asphalt, is heated by the raffinate (asphalt) heater and then fed to the extract (asphalt) flash drum, where asphalt is separated from solvent. The solvent coming out of the raffinate (asphalt) flash drum, after liquefied, is recycled to the solvent drum No. 1. A trace amount of solvent remaining in asphalt is removed by the asphalt stripper.

Conditions for extractor operation are approximately as follows:

Temperature	100–180°C
Pressure	15–28 kg/cm^2 A
Solvent volume	A ratio of 2 to 4 to the feed volume.

(3) Process yields

Estimated yields from the M-DS process are given in Table 5.4, below.

Using Orinoco vacuum residue as the feed, the M-DS process gives 68.57 vol. % (or 65.33 wt. %), based on the volume (or weight) of the feed, of deasphalted oil and 31.43 vol. % (or 34.67 wt. %) of asphalt.

Table 5.4 Yield of M-DS Process

Feed	vol. %	°API	wt. %
Orinoco			
Vacuum Residue (995°F+)	100.0	1.8	100.0
Products			
DAO	68.57	8.5	65.33
SDA Asphalt	31.43	-10.6	34.67

Properties	Vac. Residue	DAO	SDA Asphalt
°API	1.8	8.5	-10.6
Sp. Gr. 15/4°C	1.062	1.0108	1.1706
S. wt. %	4.32	3.52	5.82
Vis. @210°F, cst	–	700	
(@250°C, cp)			(4,000)
@300°F, cst	2,890	69	
(@300°C, cp)			(500)
Nitrogen, wt. %	0.82	0.418	1.58
Nickel, ppm	162	39.3	393
Vanadium, ppm	654	107.9	1,683
Asphalten (C_7), wt. %	–	0.00	43.3
Con. Carbon, wt. %	25.7	9.07	57.0
R & B Soft. pt. °C	–	–	162

5.2 PROCESS SCHEMES

This Section deals with the descriptions of process schemes wherein afore-mentioned upgrading process units are incorporated to upgrade the Orinoco heavy oil. For detailed information, refer to Sections 1.2, 2.2 and 3.2 of Volume II.

5.2.1 Fluid Coker Case

Fig. 5.4 shows a process scheme diagram for the Fluid coker case.

The feed to the refinery is a mixed crude oil consisting of a raw crude and a diluent. This mixed crude feed is fractionated at the atmospheric distillation unit into light gas, naphtha, light gas oil, heavy gas oil and atmospheric residue. Of the amount of light gas oil produced by the atmospheric distillation unit, that portion of it used to mix with crude oil is branched and recycled to the oil field for use as the diluent in oil production. The remaining portion of light gas oil and other fractions are respectively passed to subsequent process steps. The atmospheric residue from the atmospheric distillation unit is then fed to the vacuum flashing unit and fractionated into vacuum gas oil and vacuum residue.

The vacuum residue, the heaviest residue obtained from distillation, is used as the feed to the afore-mentioned various upgrading process units. The upgraded oils obtained, serving as the base oils for improved crude oil, are combined with other light fractions coming out of both the atmospheric and vacuum distillation units. A part of vacuum residue is taken out and consumed as a fuel for process heaters of the refinery.

Vacuum residue is fed to the fluid coker and cracked to give cracked gas, coker naphtha, coker gas oil and fluid coke. In addition, coke combustion in this process generates CO gas in the form of a high-temperature low-calory gas.

Naphtha, light gas oil and heavy gas oil from the atmospheric distillation unit, vacuum gas oil from the vacuum distillation unit, and coker naphtha and coker gas oil from the fluid coker process are all combined together and passed through hydrotreater to saturate those unsaturated hydrocarbons in such oil fractions and make them stable. There is thus obtained an improved crude oil having lower specific gravity and containing less sulfur than the raw crude oil.

CO gas from the fluid coker process finds its use in waste heat utilization and steam generation caused by burning this low-calory gas.

Fluid coke is employed as the boiler fuel to supply the refinery with steam, which, in turn, is used to generate power, both for refinery consumption and for raw crude production.

The surplus fluid coke, other than the amount used as boiler fuel, is stockpiled in an open area outside the refinery for future use as fuel.

The cracked gas from the fluid coker process and the gas coming out of the hydrotreater get rid of hydrogen sulfide at respective acid gas treatment units. These gases are then used as part of fuel gas, recycled hydrogen gas and the feed gas to the H_2 plant.

Fuel gas is used mainly as the process fuel and as a supplementary fuel for boiler which uses cold coke as the main fuel. The surplus amount of fuel gas is used for portable boiler to generate steam required for raw crude oil production.

Hydrogen sulfide is removed by acid gas treatment units and recovered by the Claus sulfur recovery unit. The sulfur element recovered is solidified and shipped as a by-product.

Tail gas from the sulfur recovery unit is not recovered, but is burnt into the atmosphere. Hydrogen required by the hydrotreater is produced at the H_2 plant by using natural gas supplied from outside, as the raw material. In this fluid coker case, however, part of the purge gas from the hydrotreater is used as part of the feed to the H_2 plant, after the purge gas has been desulfurized.

Two trains, in all, of units will be installed for the entire process. Periodical repair work is carried out alternately for one train at a time, so that the refinery can secure at least 50% operation and keep supplying fuel gas to the oil field and fluid coke to the refinery boiler.

5.2.2 Eureka Case

Fig. 5.5 shows a process scheme diagram for the Eureka case.

As in the above fluid coker case, the mixed crude is processed by both of atmospheric and vacuum distillation units. Fractions from these units are used as the base oils for improved crude oil.

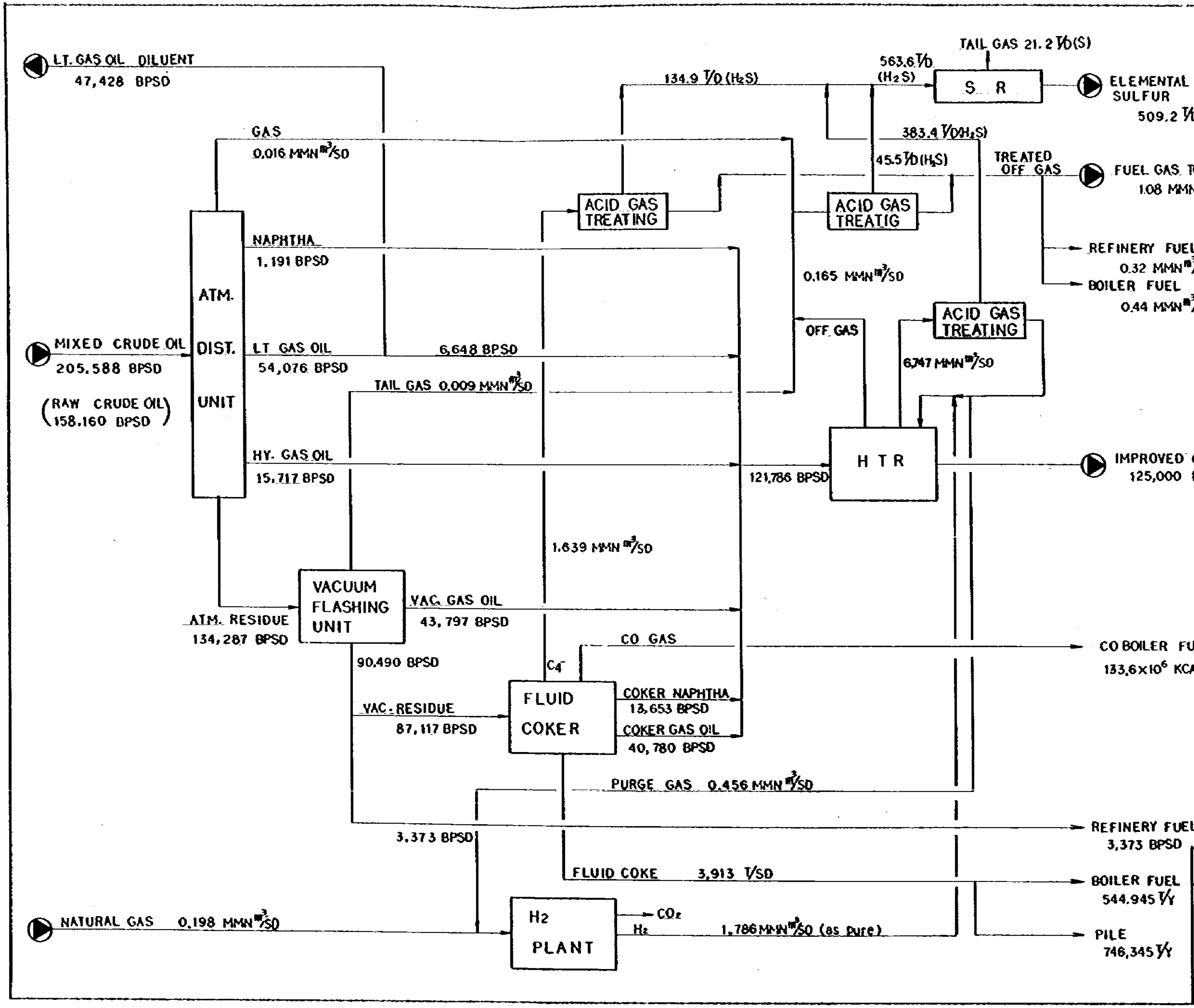
A part of vacuum residue is set aside as process fuel, and the remaining vacuum residue is fed to the Eureka process and upgraded to give cracked gas, cracked light oil, cracked heavy oil, and Eureka pitch.

The naphtha coming out of the atmospheric distillation unit is the saturated hydrocarbons containing a less amount of sulfur. It is mixed with other fractions with no further treatment.

Light gas oil, heavy gas oil and the cracked light oil from the Eureka process are combined together into a light oil fraction, which is desulfurized by the first hydrotreater.

Vacuum gas oil and the cracked heavy oil from the Eureka process is combined with each other to form a heavy oil fraction, which is desulfurized by the second hydrotreater.

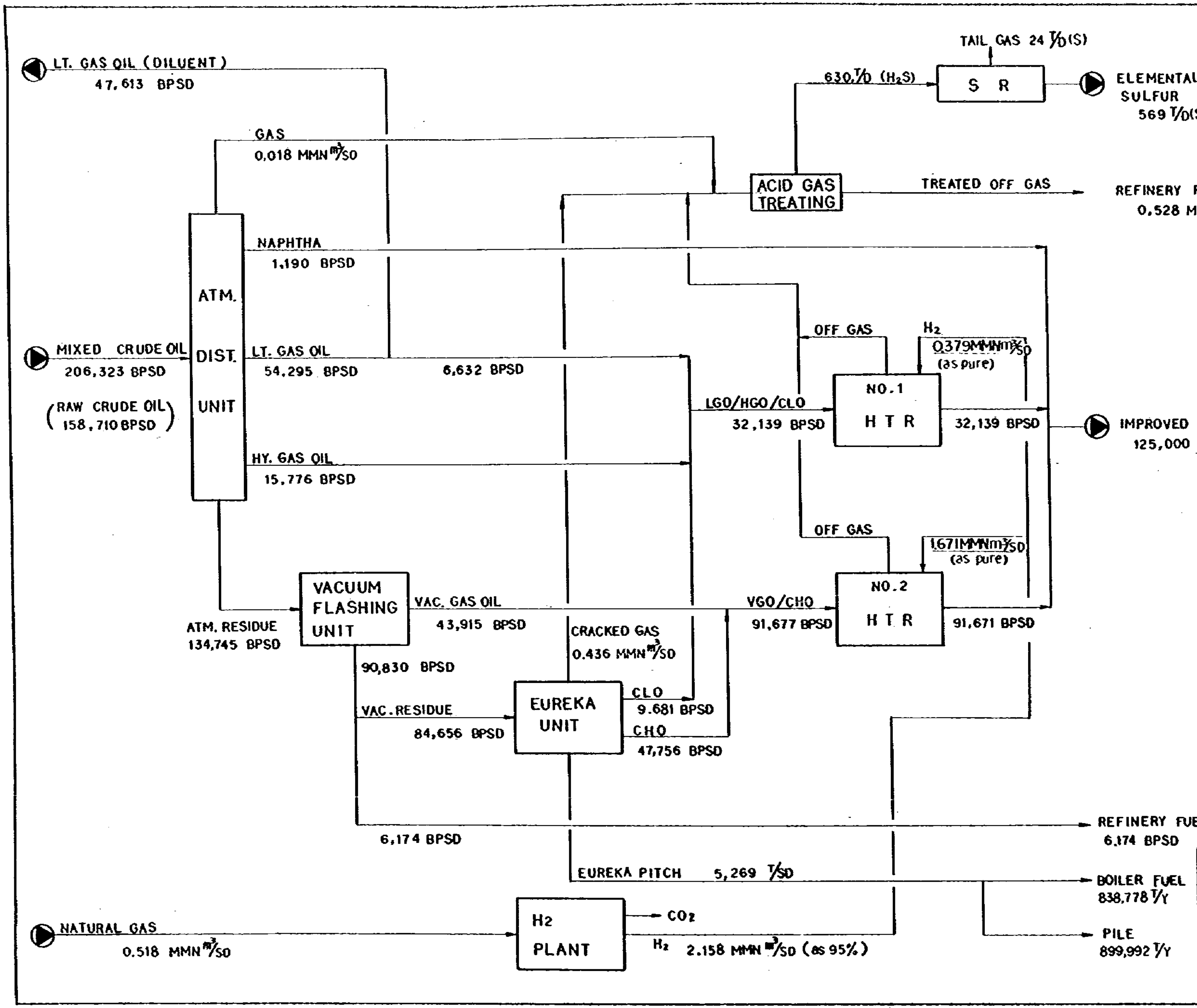
Hydrotreated light and heavy oil fractions and naphtha are combined together to give improved crude oil. The first hydrotreater treats a feed with sulfur content of 1.67 wt. % and reduces its content down to 0.1 wt. %; the second hydrotreater treats a feed with a sulfur content of 3.45 wt. % and reduces its content down to 0.5 wt. %. Therefore, the improved crude oil will have a sulfur content of 0.4 wt. %.



ORINOCO HEAVY OIL UPGRADING PROJECT

BLOCK FLOW DIAGRAM OF FLUID COKER CASE

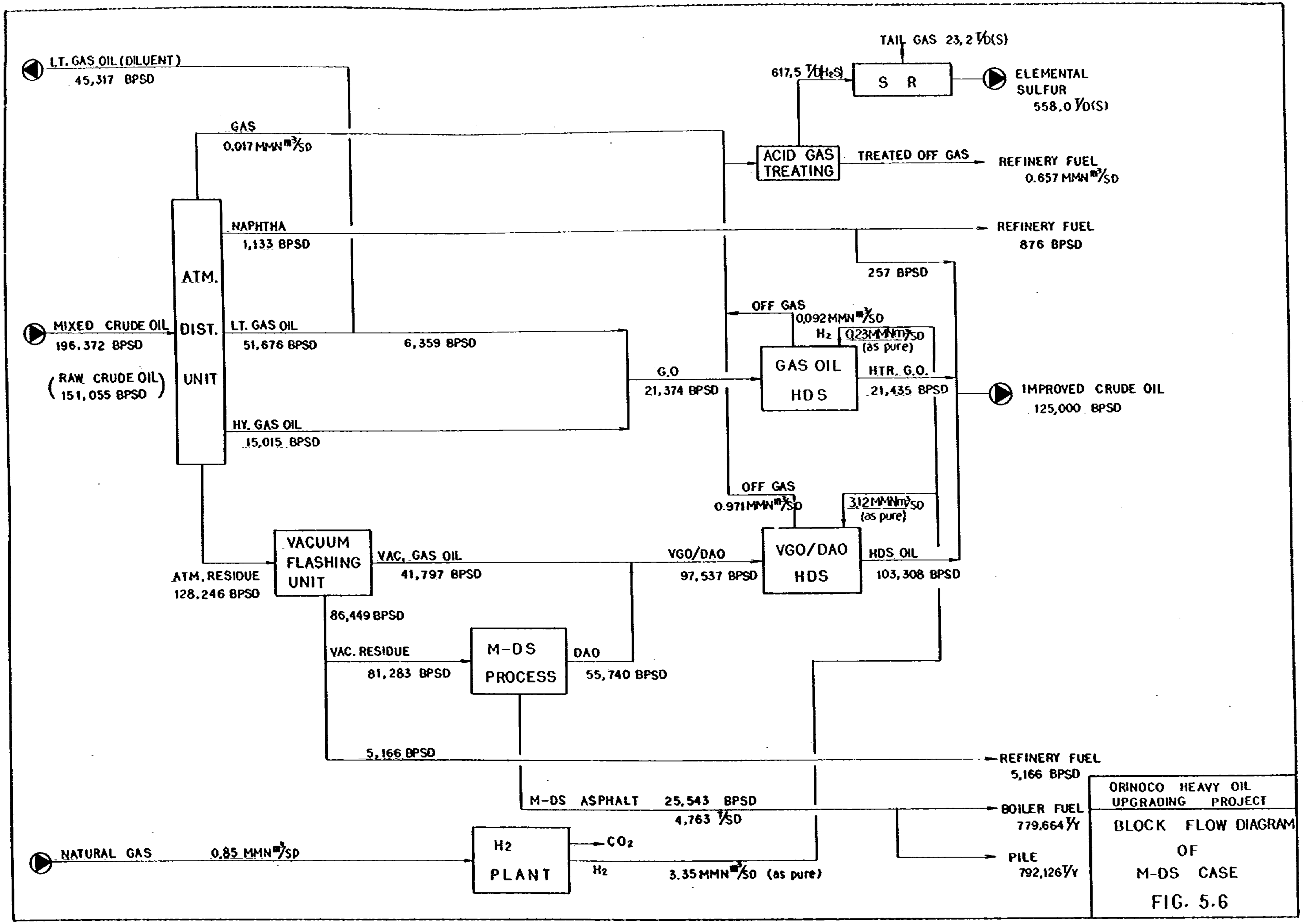
FIG. 5.4



ORINOCO HEAVY OIL UPGRADING PROJECT

BLOCK FLOW DIAGRAM OF EUREKA CASE

FIG. 5-5



ORINOCO HEAVY OIL UPGRADING PROJECT

BLOCK FLOW DIAGRAM OF M-DS CASE

FIG. 5.6

The units of acid gas treatment, sulfur recovery, and hydrogen generation used in this process are similar to those used in the Fluid coker case.

Pitch from the Eureka process is used as boiler fuel. The surplus amount is stockpiled in an open area. Fuel gas, after treated by the acid gas treating unit, is consumed entirely as process fuel within the refinery. Therefore, no fuel gas can be supplied to the oil field.

The Eureka process, too, is planned to have two trains of process units for the entire process.

5.2.3 M-DS Case

Fig. 5.6 shows a process scheme diagram for the M-DS case.

Like the above two cases, the M-DS case utilizes pre-treatment by means of both the atmospheric and vacuum distillation units.

Vacuum residue is subjected to M-DS process steps, in which the solvent deasphalting process separates the feed into deasphalted oil and asphalt.

Light gas oil and heavy gas oil from the atmospheric distillation unit are treated by gas oil hydrodesulfurizer for the only purpose of desulfurization. This treatment for relatively easy desulfurization reduces the sulfur content from 2.06 wt. % to 0.15 wt. %.

Vacuum gas oil and the deasphalted oil from the M-DS process are combined and treated by VGO/DAO hydrodesulfurizer. This process is used under severe operating conditions because DAO is not only heavy, but also contains relatively large amounts of metals.

Furthermore, the improved crude oil is required to have a specific gravity as light as about 25° API. In order to meet this requirement, this process has to be provided with a hydrocracking function. As a result, this process is operated under conditions as severe as in directdesulfurization, to give a product of 25.2° API by using a feed of 10.8° API. Keeping step with severe operating conditions, the rate of desulfurization becomes higher, affording sulfur reduction down to 0.03 wt. %.

Naphtha from atmospheric distillation is partly used as fuel for the H_2 plant. The remainder is mixed with desulfurized oils to give an improved crude oil with a sulfur content of 0.05 wt. %.

Acid gas treating unit, sulfur recovery unit and H_2 plant used in the M-DS case are similar to those used in the above two cases. It should be noted, however, that the upgrading process of this case does not involve a cracking process. Gas evolves from hydrodesulfurizers only, and there is hardly obtained any surplus amount of gas. The whole amount of gas is thus consumed within the refinery. In spite of this, fuel is still in shortage, and therefore, naphtha is used as the fuel for the H_2 plant. It is impossible in this case to supply fuel gas to the oil field.

As described in more details in Chapter 7, the asphalt obtained from the M-DS process is used as boiler fuel in its high-temperature liquid state, without storage. The surplus amount of asphalt is taken out of the refinery, in its high-temperature liquid state. This case, too, uses two trains of units.

Tables 5.5, 5.6, 5.7, 5.8 and 5.9 summarize overall material balance, properties of improved crude oils, capacities of process units, hydrogen balance and sulfur balance for all the three cases.

Table 5.5 Overall Material Balance

Normal Operation Case

	Case	Fluid Coker	Eureka	M-DS
Feed				
Mixed Crude	BPSD	205,588	206,323	196,372
Natural Gas	MMNm^3/SD	0.198	0.518	0.850
Products				
Improved Crude	BPSD	125,000	125,000	125,000
Lt. Gas Oil (Diluent)	BPSD	47,428	47,613	45,317
Elemental Sulfur	T/SD	509.2	569.0	558.0
Fuel Gas** (To Oil Field)	MMNm^3/SD (MMKcal/H)	1.08 (563)	0	0
Electric Power*	MW	126.2	126.6	120.5
Gypsum*	T/SD	673	498	594
Surplus By-product	T/Y	746,345	899,992	792,126
Oil Production				
Raw Crude Oil	BPSD	158,160	158,710	151,055
Raw Crude Oil	BPCD	142,994	143,491	136,570
Electric Power (Base: 150 MW/170,000 BPCD)	MW	126.2	126.6	120.5

* 365 D/Y production

** 0.485 MMNm^3/SD in case of process 1 train operation

Table 5.6 Properties of Improved Crude Oils

Case	Fluid Coker	Eureka	M-DS
Feed			
(Raw Crude Oil)			
°API	8.5	8.5	8.5
Sulfur, wt. %	3.67	3.67	3.67
Products			
(Improved Crude Oil)			
°API	25.7	25.0	26.1
Sulfur, wt. %	0.70	0.41	0.05
Viscosity			
cst. @ 100°F	4.9	–	22.0
cst. @ 210°F	2.8	–	4.0
Nitrogen, wt. %	0.17	–	0.008
CCR, wt. %	0.147	–	0.13
Components, vol. %			
SR. Naphtha	–	1.0	0.2
HTR. SR. Naphtha	0.9	–	–
HTR. SR. LGO	5.5	5.3	5.1
HTR. SR. HGO	12.9	12.6	12.1
HTR. VGO	36.0	35.1	35.4
HTR. Coker Naphtha	11.2	–	–
HTR. Coker Gas Oil	33.5	–	–
HTR. Cracked Light Oil	–	7.8	–
HTR. Cracked Heavy Oil	–	38.2	–
HTR. DAO	–	–	47.2
Yield of Distillation			
C_5/375°F, vol. %	15.0	7.3	9.5
375/650°F, vol. %	30.0	32.4	34.0
650/1,000°F, vol. %	55.0	60.3	33.5
1,000°F+, vol. %	–	–	23.0
Sulfur Content of Distillation			
C_5/375°F, wt. %	0.24	0.09	0.01
375/650°F, wt. %	0.67	0.1	0.08
650/1,000°F, wt. %	0.73	0.6	0.02
1,000°F+, wt. %	–	–	0.03

Table 5.7 Capacities of Process Units

Case		Fluid coker		Eureka		M-DS	
Process Units	Unit	Capacity	No's	Capacity	No's	Capacity	No's
Atmospheric Distillation	BPSD	102,800	2	103,200	2	98,200	2
Vacuum Flashing	BPSD	67,200	2	67,400	2	64,200	2
Fluid Coker	BPSD	43,600	2	–		–	
Eureka	BPSD	–		42,400	2	–	
M-DS	BPSD	–		–		40,700	2
Hydrotreater	BPSD	60,900	2	16,100	2	10,700	2
	BPSD	–		45,900	2	48,800	2
Hydrogen Generation	$MMNm^3/D$ as H_2	0.90	2	1.08	2	1.70	2
Acid Gas Treating	T/SD as H_2S	67 23 192	2 2 2	315.0 – –	2	309 – –	2
Sulfur Recovery	T/SD as S	255	2	285	2	279	2

Table 5.8 Hydrogen Balance

Case	Fluid Coker	Eureka	M-DS
Feed of H_2 Plant			
Natural Gas to H_2 plant	0.198 $MMNm^3/D$	0.518 $MMNm^3/D$	0.850 $MMNm^3/D$
Purge Gas to H_2 plant	0.456 $MMNm^3/D$	–	–
Generated Hydrogen			
Hydrogen from H_2 plant (as pure)	1.786 $MMNm^3/D$	2.050 $MMNm^3/D$	3.350 $MMNm^3/D$
Consumption of Hydrogen (as pure)			
Hydrotreater	1.786 $MMNm^3/D$	–	–
No. 1 Hydrotreater	–	0.379 $MMNm^3/D$	–
No. 2 Hydrotreater	–	1.671 $MMNm^3/D$	–
Gas Oil HDS	–	–	0.230 $MMNm^3/D$
VGO/DAO HDS	–	–	3.120 $MMNm^3/D$
Consumption Rate of Hydrogen			
Hydrotreater	547 SCF/B	–	–
No. 1 Hydrotreater	–	440 SCF/B	–
No. 2 Hydrotreater	–	680 SCF/B	–
Gas Oil HDS	–	–	401 SCF/B
VGO/DAO HDS	–	–	1,193 SCF/B

Table 5.9 Sulfur Balance
Contained Sulfur (T/SD)

Normal Operation Case

Case		Fluid Coker	Eureka	M-DS
– Input –				
Raw Crude Oil (3.67 wt.%S)		932.7	936.0	890.8
– Output –				
Improved Crude Oil		125.2	71.9	8.9
(Fluid Coker	0.7 wt.%S)			
(Eureka	0.4 wt.%S)			
(M-DS	0.05 wt.%S)			
Product Elemental Sulfur		509.2	569.0	558.0
Tail Gas from Sulfur Recovery Unit		21.2	24.0	23.2
Flue Gas from Process Furnace (Vacuum Residue & Naphtha Fuel)		24.6	45.0	38.5
(Produced Total By-product)		(252.5)	(226.1)	(262.2)
Fluid Coke	5.79 wt.%S			
Eureka pitch	4.3 wt.%S			
M-DS Asphalt	5.82 wt.%S			
CO gas				
Gypsum		125.3	92.7	110.5
Tail gas from Flue gas Desulfurization		13.9	10.3	12.3
Surplus By-product		113.3	123.1	139.4

Attachment to Table 5.6 Properties of Improved Crude Oil on Fluid Coker Case

Fractions	Light Naphtha	Heavy Naphtha	Keresene	Gas Oil	Vacuum Gas Oil	Residue	Improved Crude Oil
Properties							
TBP Fraction, °F	C_5–375	–	–	375–650	650–1000	–	C_5–1000
Yield, vol %	15.0	–	–	30.0	55.0	–	100
Gravity, °API	47.5	–	–	27.4	20.4	–	25.7
Sulfur, wt%	0.24	–	–	0.67	0.73	–	0.70
Nitrogen, ppm	–	–	–	1080	2400	–	1700
RON, Clear	–	–	–	–	–	–	–
PONA, vol %	–	–	–	–	–	–	–
Smoke pt. mm	–	–	–	–	–	–	–
Cetane Ix.	–	–	–	–	–	–	–
Diesel Ix.	–	–	–	–	–	–	–
CCR, wt%	–	–	–	–	0.26	–	0.147
Aniline pt. °F	–	–	–	–	–	–	–
Bromine No. gBr/100mg	–	–	–	0.1>	0.1>	–	0.1>
Metal Content							
V, ppm	–	–	–	–	0.22	–	0 2
Ni, ppm	–	–	–	–	0.12	–	0.1
Viscosity							
cst @ 100°F	–	–	–	–	–	–	4.9
cst @ 210°F	–	–	–	1.0	4.0	–	2.8

Attachment to Table 5.6 Properties of Improved Crude Oil on Eureka Case

Fractions	Light Naphtha	Heavy Naphtha	Keresene	Gas Oil	Vacuum Gas Oil	Residue	Improved Crude Oil
Properties							
TBP Fraction, °F	C_5–236	236–344	344–500	500–666	666–965	–	C_5–965
Yield, vol %	3.48	3.82	10.42	22.27	60.01	–	100
Gravity, °API	70	49.9	39.3	24.2	20.1	–	25
Sulfur, wt%	0.06	0.09	0.16	0.25	0.54	–	0.41
Nitrogen, ppm	10	40	150	200	600	–	–
RON, Clear	–	–	–	–	–	–	–
PONA, vol %	–	–	–	–	–	–	–
Smoke pt. mm	–	–	23	–	–	–	–
Cetane Ix.	–	–	–	38–40	–	–	–
Diesel Ix.	–	–	–	–	–	–	–
CCR, wt%	–	–	–	–	0.58	–	–
Aniline pt. °F	–	–	–	–	–	–	–
Bromine No. gBr/100mg	15	14	11	–	–	–	–
Metal Content							
V, ppm	–	–	–	–	0.09	–	–
Ni, ppm	–	–	–	–	0.30	–	–
Viscosity							
cst @ 100°F	–	–	–	–	–	–	–
cst @ 210°F	–	–	–	–	–	–	–

Attachment to Table 5.6 Properties of Improved Crude Oil on M-DS Case

Fractions	Light Naphtha	Heavy Naphtha	Keresene	Gas Oil	Vacuum Gas Oil	Residue	Improved Crude Oil
Properties							
TBP Fraction, °F	C_5–375	–	–	375–650	650–1000	1000^+	C_5^+
Yield, vol %	9.5	–	–	34.0	33.5	23.0	100
Gravity, °API	55.0	–	–	32.5	22.9	9.0	26.1
Sulfur, wt%	0.01	–	–	0.08	0.02	0.03	0.05
Nitrogen, ppm	10	–	–	45	90	135	80
RON, Clear	60	–	–	–	–	–	–
PONA, vol %	–	–	–	–	–	–	–
Smoke pt. mm	–	–	–	–	–	–	–
Cetane Ix.	–	–	–	45	–	–	–
Diesel Ix.	–	–	–	52	41	–	–
CCR, wt%	–	–	–	–	0.10	0.39	0.13
Aniline pt. °F	–	–	–	160	180	–	–
Bromine No. gBr/100mg	0.0	–	–	–	–	–	–
Metal Content							
V, ppm	–	–	–	–	0.0	0.7	0.2
Ni, ppm	–	–	–	–	0.0	0.3	0.1
Viscosity							
cst @ 100°F	–	–	–	3.4	22	4500	22
cst @ 210°F	–	–	–	–	–	80	4

CHAPTER 6 DESCRIPTION OF UTILITY AND OFFSITE FACILITIES

This chapter outlines the utility and offsite facilities of Orinoco heavy oil upgrading refinery. They include all the facilities other than the process units of the refinery. Boilers using by-products as the fuel will be separately dealt with in Chapter 7. Flue gas desulfurization unit for the treatment of boiler flue gas is described in this chapter as one of offsite facilities. Description of this chapter is based on the study results given for the three cases in Chater 4 of Volume II.

6.1 UTILITY FACILITIES

These facilities serve to generate utilities necessary to operate process units and offsite facilities of the upgrading refinery. With an exception of raw water, it is intended that all the utilities are supplied by own generation, to secure stable supplies. As a special case of refinery, the utility facilities of this refinery is planned to supply the oil field with power required for crude production.

6.1.1 Utilities Supply System

Fig. 6.1 shows the entire system for the supplies of steam, power and water. The system summarized in this figure is common to all of the 3 cases. As regards steam, the following four levels of steam are used in the refinery:

Ultra-high-pressure steam	100 kg/cm^{2}G,	500°C
High-pressure steam	50 kg/cm^{2}G,	405°C
Medium-pressure steam	16 kg/cm^{2}G,	275°C
Low-pressure steam	4 kg/cm^{2}G,	165°C

Ultra-high-pressure steam is aimed at the generation of power required for crude oil production. Contemplated for this purpose are those boilers using, as fuel, a heavy residual by-produced from either one of the three upgrading processes.

High-, medium- and low-pressure steam are used for steam balance and power balance among process units and other facilities in the refinery. In the Fluid coker case, however, CO gas can be utilized to generate high-pressure steam by means of CO boilers. Both of condensation turbines and extraction turbines are used for power generation based on above types of steam. Steam-condensed water is recovered, combined with additional amounts of fresh boiler feed water, and recycled to boilers.

Cooling water for use by process units and for steam condensation is recycled, thus requiring a cooling tower.

Refinery off-gas and vacuum residue are used as fuel, but in the case of M-DS process, off-gas produced is not sufficient to cover the whole fuel needs of the refinery. In this case, therefore, naphtha is partly used as the fuel for H_2 plant.

In addition, compressed air is produced for instrumentation and process air use. Nitrogen gas is produced by air separation and used as an inert gas.

6.1.2 Utilities Balances

Tables 6.1, 6.2 and 6.3 give the utilities balance for each upgrading case. These tables indicate consumption and production of utilities, such as power, each pressure level of steam, boiler feed water, condensed water, process injection water, foul water, loss water, process cooling water, mechanical cooling water, and fuel. In the table, positive values indicate consumption, and negative values production. The utility requirements of the refinery are summarized in Table 6.4.

6.1.3 Capacities of Utility Facilities

Table 6.5 gives the capacities and number of units (including the number of stand-by units) required to maintain the afore-mentioned utilities balance.

6.2 OFFSITE FACILITIES

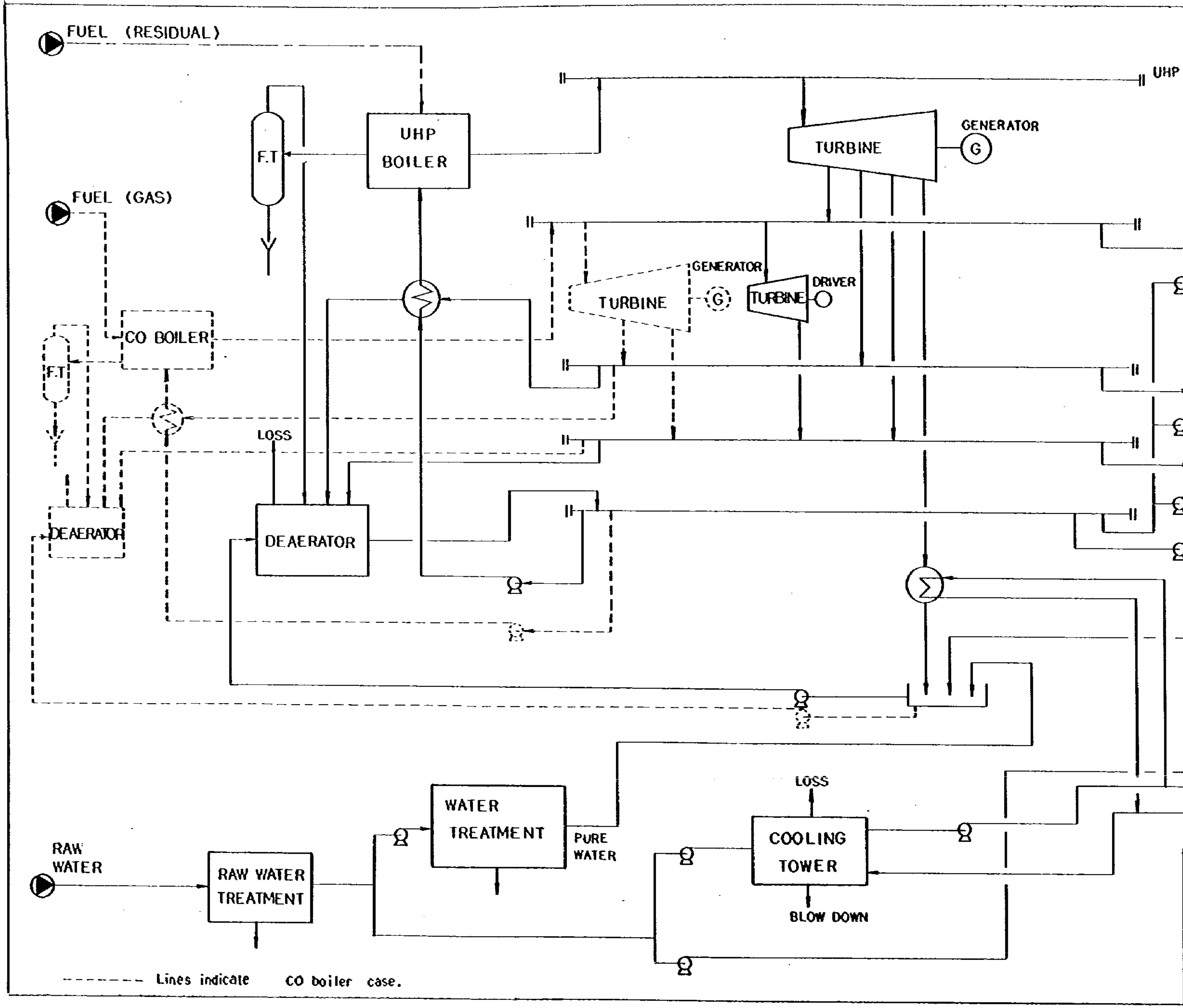
The offsite facilities described in this Section are understood to include all the necessary facilities in the refinery other than process units and utility facilities.

6.2.1 Storage Facilities

Oil handling system of this refinery is as shown in the tank flow diagram of Fig. 6.2. The system illustrated in this diagram is common to all the three cases. Mixed crude oil, fed to the refinery as its raw material, is stored in the refinery in an amount sufficient to operate the refinery for a month, so that stable operation can be secured without being affected by any change in the crude production situations. Since the refinery is close to the oil field, with a pipeline laid between them, such an amount of crude oil can also be stored at the main crude processing station in the oil field. From refinery standpoint, however, storage within the refinery is considered more convenient.

Intermediate tanks are not installed fully, assuming that two complete trains of process units are used, that process units are started up or shut down, with one train at a time, and that oils are charged to processes under a hot charge system. Along this line of policy, only the upgrading process and the hydrodesulfurization process are provided with feed tanks capable of storing respective feeds in amounts sufficient to run these processes for a week.

As for the storage of products, consisting of improved crude oil and light gas oil used as diluent, it is planned that improved crude oil is stored within the refinery in an amount corresponding to a week of production, assuming that it is pipelined to the shipping port and that a tank site is located alongside the shipping port for the con-



ORINOCO HEAVY OIL UPGRADING PROJECT

STEAM, POWER AND WATER SYSTEM DIAGRAM

FIG. 6.1

Table 6.1 Utility Balance of Fluid Coker Case

Normal Operation Case

	Elec.	STEAM						WATER								FUEL	
	Power	UHP	HP	HP (sat)	MP	LP	Loss	BFW	Cond.	Pure Water	Indust	Raw	Foul	Cooling	Mecha.	Cons.	Gen.
Item	KW	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	MMKcal/H	
Process Units (1 train)	16,420	0	52.6	-40.8	22.7	91.5	-80.1	237.2	-136.2	0	34.1	0	-181	4,610	125	193	-1,311.5
Process Units (1 train)	16,420	0	52.6	-40.8	22.7	91.5	-80.1	237.2	-136.2	0	34.1	0	-181	4,610	125	193	-1,311.5
CO Boiler/Generator (1 train)	-16,590	0	-52.6	40.8	-22.7	-91.5	-2.2	-237.2	136.2	231.2	0	0	-2	0	47	149	0
CO Boiler/Generator (1 train)	-16,590	0	-52.6	40.8	-22.7	-91.5	-2.2	-237.2	136.2	231.2	0	0	-2	0	47	149	0
Offsite Facility	4,300	0	0	0	10.2	61.5	-15	0	-61.7	0	5	0	0	0	9	0	0
By-product Utilization (Boiler/Generator/FGD)	-148,930	0	0	0	-10.2	-61.5	-104	0	61.7	21.8	280	0	-187.8	24,790	528	559	0
Utility Facility	16,570	0	0	0	0	0	-1,024	0	0	-484.2	-353.2	2,112.4	-251	-34,010	-881	3	0
Total	-128,400	0	0	0	0	0	-1,307.6	0	0	0	0	2,112.4	-804.8	0	0	1,246	-2,623

Power: To Field 126,200
Excess 2,200

Loss: Loss

Raw: Raw Water Make-up

Foul: Disposal

Fuel (Cons. + Gen.): -1,377

Fuel gas to Field 564
Coke to pile 813

Table 6.2 Utility Balance of Eureka Case

Normal Operation Case

Item	Elec. Power	STEAM UHP	HP	HP (sat)	MP	LP	Loss	WATER BFW	Cond.	Pure Water	Indust	Raw	Foul	Cooling	Mecha.	FUEL Cons.	Gen.
	KW	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	MMKcal/H	
Process Unit (1 train)	16,830	0	14.1	0	80.2	77.0	-65.1	290.1	-167.8	0	35.6	0	-264.1	9,490	142	359	-1,349
Process Unit (1 train)	16,830	0	14.1	0	80.2	77.0	-65.1	290.1	-167.8	0	35.6	0	-264.1	9,490	142	359	-1,349
Offsite Facility	4,300	0	0	0	11.7	61.7	-15.0	0	-63.4	0	5.0	0	0	0	10	0	0
By-product Utilization (Boiler/Generator/FGD)	-180,100	0	-28.2	0	-172.1	-215.7	-99.9	-580.2	399	612.9	185.0	0	-100.8	19,430	523	896	0
Utility Facility	15,190	0	0	0	0	0	-1,150	0	0	-612.9	-261.2	2,310.1	-286	-38,410	-817	4	0
Total	-126,950	0	0	0	0	0	-1,395.1	0	0	0	0	2,310.1	-915	0	0	1,618	-2,698
	↓ To Field 126,600 Excess 350						↓ Loss					↓ Raw water Make-up	↓ Disposal			-1,080 ↓ pitch to pile	

Table 6.3 Utility Balance of M-DS Case

Normal Operation Case

	Elec.	STEAM						WATER								FUEL	
	Power	UHP	HP	HP (sat)	MP	LP	Loss	BFW	Cond.	Pure Water	Indust	Raw	Foul	Cooling	Mecha.	Cons.	Gen.
Item	KW	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	T/H	MMKcal/H	
Process Unit (1 train)	21,440	0	24.7	0	58.2	125.9	-62.2	178.7	-216.7	0	83.9	0	-192.5	6,190	127	312	-1,242.5
Process Unit (1 train)	21,440	0	24.7	0	58.2	125.9	-62.2	178.7	-216.7	0	83.9	0	-192.5	6,190	127	312	-1,242.5
Offsite Facility	4,300	0	0	0	10.7	57.5	-14.0	0	-59.2	0	5.0	0	0	0	9	0	0
By-product Utilization (Boiler/Generation/FGD)	-180,800	0	-49.4	0	-127.1	-309.3	-101.4	-357.4	492.6	383.8	233	0	-164.8	17,470	508	870	0
Utility Facility	12,200	0	0	0	0	0	-897	0	0	-383.8	-405.8	1,906.6	-220	-29,805	-771	2	0
Total	-121,420	0	0	0	0	0	-1,136.8	0	0	0	0	1,906.6	-769.8	0	0	1,496	-2,485
	↓ To Field 120,500 Excess 920						↓ Loss					↓ Raw water Make-up	↓ Disposal			↓ -989 ↓ Asphalt to Pile	

Table 6.4 Summary of Utility Requirements

			Normal Operation Case	
Requirement	Unit	Fluid Coker	Eureka	M-DS
Electric Power				
for Oil Field	MW	126.2	126.6	120.5
for Refinery	MW	74.2	81.4	82.5
Steam				
Ultra High Pressure	T/H	772.1	1,190.3	1,180.4
High Pressure	T/H	388.4	–	–
Cooling Water (Circulation)				
Process Cooling	T/H	9,770	19,490	12,830
Surface Condenser	T/H	24,600	19,300	17,300
Mechanical Cooling	T/H	930	870	815
Net Boiler Feed Water	T/H	484	613	384
Net Raw Water Intake	T/H	2,112	2,310	1,907
Fuel				
Liquid	MM kcal/H	220	410	394
Gas	MM kcal/H	523	308	230
Residual	MM kcal/H	503	900	872

Table 6.5 Capacities of Utility Facilities

Facility	Fluid Coker Case		Eureka Case		M-DS Case		Note
	Capacity	No.s	Capacity	No.s	Capacity	No.s	
1. Steam Generator							
Ultra High Pressure Steam	260 T/H	4	240 T/H	6	240 T/H	6	one unit for stand-by
High Pressure Steam	200 T/H	2	–		–		
2. Power Generator							
by Ultra High Pressure Steam	55,000 KW	4	46,000 KW	6	44,000 KW	6	one unit for stand-by
by High Pressure Steam	18,000 KW	2	–		–		
3. BFW Treatment							
Activated carbon Adsorption	250 T/H	3	310 T/H	3	200 T/H	3	one unit for stand-by
Ion Exchange Demineralization	250 T/H	3	310 T/H	3	200 T/H	3	one unit for stand-by
Condensate Tank	9,000 KL	2	10,000 KL	2	10,500 KL	2	
BFW Tank	6,000 KL	2	7,500 KL	2	4,600 KL	2	
4. Cooling Water System							
Cooling Tower	18,000 T/H	2	20,000 T/H	2	15,500 T/H	2	
Raw Water Tank	25,000 KL	2	28,000 KL	2	23,000 KL	2	
5. Air System	1,800 Nm^3/H	3	2,000 Nm^3/H	3	1,900 Nm^3/H	3	one unit for stand-by
6. Inert Gas System	350 Nm^3/H	2	350 Nm^3/H	2	350 Nm^3/H	2	
7. Potable Water System							
Chlorinator	5 T/H	1	5 T/H	1	5 T/H	1	
Tank	500 KL	1	500 KL	1	500 KL	1	
Elevated Tank	10 KL	1	10 KL	1	10 KL	1	

Table 6.6 Tank Summary

Case		Fluid Coker		Eureka		M-DS	
Tank capacity & Number		KL	No.s	KL	No.s	KL	No.s
Mixed Crude	FR	133,000	8	133,000	8	127,000	8
Diluent LGO	CR	28,000	2	28,000	2	26,500	2
Improved Crude	FR	75,000	2	75,000	2	75,000	2
Vacuum Residue	CR	51,000	1	50,000	1	48,000	1
HTR Feed	CR	–		19,000	1	13,000	1
HTR Feed	CR	62,000	1	54,000	1	56,000	1
Oily Slop	CR	4,000	1	4,000	1	4,000	1
Fuel Naphtha	DR	–		–		500	1
HTR Feed Naphtha	FR	9,000	1	–		–	
Others	CR	40,000	–	–		–	
Total		1,436,000		1,397,000		1,340,500	

FR: Floating Roof

CR: Cone Roof

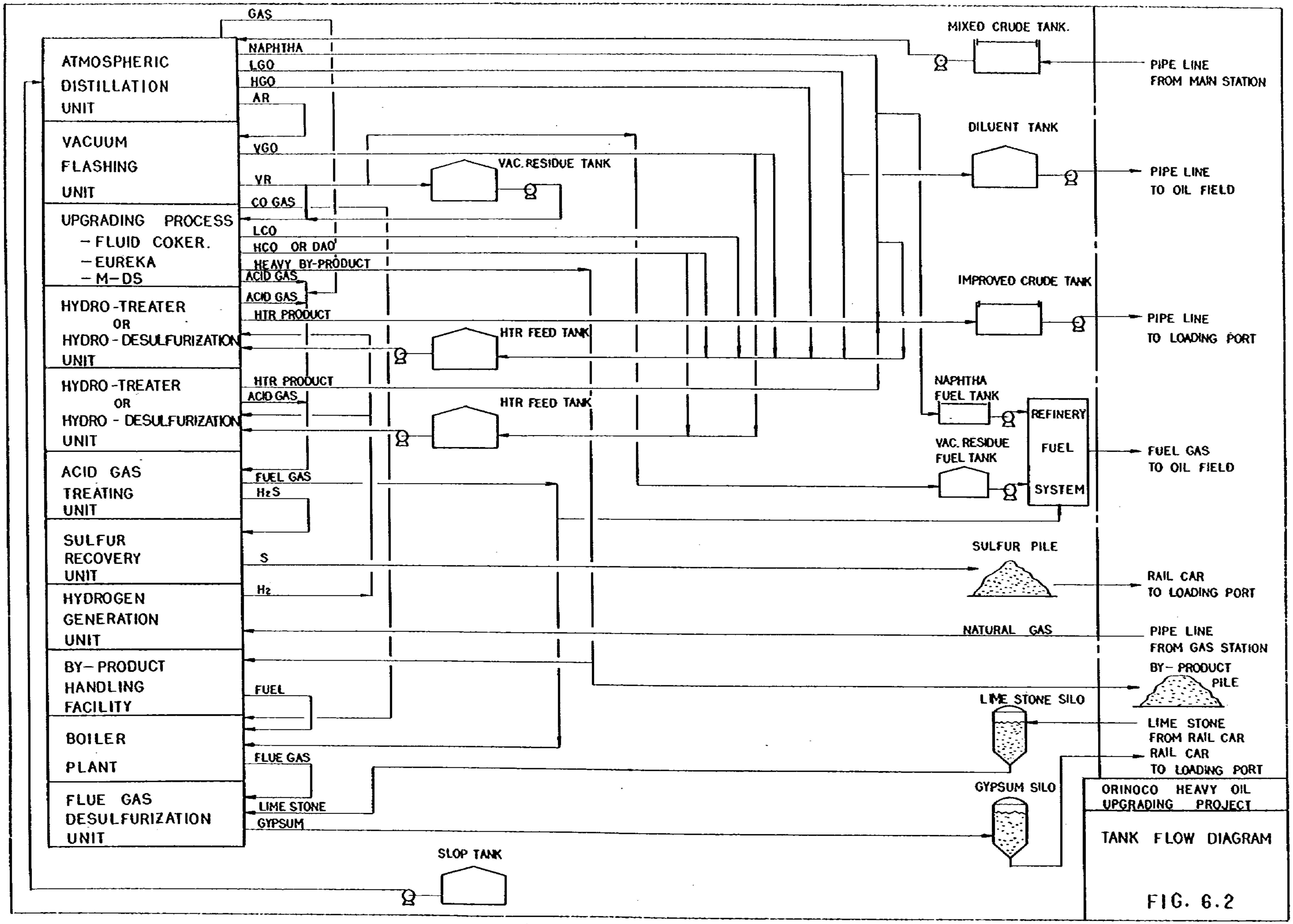
DR: Dome Roof

This summary does not include tanks for water.

venience of tanker loading. In the case of diluent storage, it is condidered more convenient to locate a diluent tank in the oil field and control the storage volume in response to crude oil production. It is thus planned that the diluent tank within the refinery, connected to the tank in the oil field by a pipeline, would store diluent in an amount corresponding to a week of production. Table 6.6 gives the tank summary of each case.

6.2.2 Product Shipping System

The product of improved crude oil is transported from the refinery product tank to the terminal of the shipping port by way of a pipeline. Now that the shipping port is not yet determined under this project, conditions can not be fixed for the product transport through the pipeline. Therefore, product shipping plan is limited up to the stage where improved crude oil is pumped out of the refinery at a discharge pressure of 40



ORINOCO HEAVY OIL UPGRADING PROJECT

TANK FLOW DIAGRAM

FIG. 6.2

kg/cm^2G. This report thus excludes the study on the pipeline and a booster pump station, if such a station is deemed necessary.

Diluent has to be sent to the tank in the oil field. Now that the site and other conditions are not yet determined under the project, the diluent shipping plan is limited up to the stage where diluent is pumped out of the refinery at a discharge pressure of 10 kg/cm^2G.

6.2.3 Solids Handling System

(1) Solid sulfur

Sulfur obtained from the sulfur recovery unit is pelletized and stockpiled in an outdoor storage yard within the refinery. This amount of storage corresponds to a week of production. Any surplus amount is transported to the shipping port and stored there in a required amount. The sulfur stored in the refinery is thus handed over to the transport facilities, study of which is not included here.

(2) Gypsum

Gypsum produced at the flue gas desulfurization unit is stored in silos in an amount corresponding to a week of production. Like sulfur, any surplus amount is handed over to the transport facilities.

(3) Limestone

Limestone, an auxiliary feed to the flue gas desulfurization unit, is stored in an amount sufficient to consume in a month. The refinery is equipped with limestone storage silos, but this study excludes consideration for the equipment required in the stages prior to the receiving of limestone into the refinery.

(4) By-products (heavy residuals) from upgrading processes

Coke and pitch from the fluid coker process and the Eureka process, respectively, are stock-piled in an outdoor storage yard outside the refinery, after boiler fuel requirement has been taken aside. This study will consider the use of belt conveyor to bring the by-product up to the fence of the refinery. Table 6.7 gives the solids storage facilities in the refinery.

Table 6.7 Storage Capacity of Solid Material

Normal Operation Case

	Case	Fluid Coker		Eureka		M-DS	
(1)	Sulfur						
	Production	509.2	T/SD	569	T/SD	558	T/SD
	Storage yard	1,200	m^2	1,300	m^2	1,300	m^2
(2)	Gypsum						
	Production	673	T/SD	498	T/SD	594	T/SD
	Silo	2,500	Tx2	1,750	Tx2	2,100	Tx2
(3)	Limestone						
	Consumption	392	T/SD	290	T/SD	346	T/SD
	Silo	4,000	Tx3	2,900	Tx3	3,500	Tx3
(4)	By-product (Residuals)	No		No		No	

6.2.4 Wastewater Treatment Facilities

Wastewater discharged from the refinery is categorized into process wastewater, oily wastewater, and clean wastewater. The first two groups of water are respectively treated before they are discharged from the refinery.

Process wastewater contains H_2S and NH_3. These components are removed by a foul water stripper. Water is then treated with a CPI oil separator. Oily water is also treated by the CPI oil separator to remove oil. Clean wastewater, such as blowdown from cooling towers and oil-free rainwater, is combined with the above treated water, and passed into guard basin, from which wastewater is discharged outside the refinery. Further facilities for the secondary and tertiary treatment is not provided under this project, because wastewater discharge standards have not yet been set. Table 6.8 gives properties of wastewater discharged after the above treatment. If wastewater properties have to be lower than these levels, treatment will require additional facilities such as active sludge treatment or activated carbon treatment facilities.

Table 6.8 Properties of Treated Wastewater

Normal operation case

Properties \ Case		Fluid Coker	Eureka	M-DS
Quantity of waste water T/H		805	915	770
Properties				
pH		6–8	6–8	6–8
H_2S	wt.ppm	2	4	3
NH_3	wt.ppm	13	22	19
SS	wt.ppm	20	16	20
COD	wt.ppm	171	208	187
Oil	wt.ppm	5	6	5

6.2.5 Flue Gas Desulfurization Unit

A large amount of SO_2 exists in the flue gas exhausted from those boilers using, as fuel, a heavy residual by-produced from upgrading processes. Desulfurization equipment is thus required as an offsite facility to prevent air pollution.

(1) Flue gas desulfurization method

There are various methods of recovering SO_2 from flue gases. In this study there has been adopted a method in which SO_2 is recovered in the form of gypsum. Flue gas coming out of the boilers goes to the scrubber, where flue gas is moistened and cooled to a predetermined temperature, and at the same time, the gas gets rid of dust in the scrubber. Flue gas is then passed into an absorbing column, in which the gas comes in contact with a calcium-containing absorbing solution, thereby getting rid of sulfur oxides in the flue gas. The desulfurized flue gas is passed through an eliminator to remove mist, and then is discharged into the atmosphere. Meanwhile, waste liquid discharged from the cooling section is treated at the cooling wastewater treatment section to separate solids from liquid.

Solids are taken out in the form of dehydrated cake. Water, after partly purged to the outside of the cooling system, is recycled to the cooling section to supplement fresh supply of cooling water.

Absorbing wastewater discharged from the absorbing section is sent to the gypsum production section, where gypsum is produced as a by-product.

(2) Material balance

Table 6.9 gives the material balance in the flue gas desulfurization for all the three cases.

Table 6.9 Material Balance of Flue Gas Desulfurization

Normal operation case

Case	Fluid Coker	Eureka	M-DS
Sulfur in Flue Gas	139.2 T/SD	103 T/SD	122.8 T/SD
Limestone as feed	392 T/SD	290 T/SD	346 T/SD
Gypsum as product	673 T/SD	498 T/SD	594 T/SD
Unrecovered Sulfur in Flue Gas	13.9 T/SD	10.3 T/SD	12.3 T/SD

6.2.6 Fire-fighting System

The refinery is equipped with the fire-preventing and fire-fighting means. Water for fire-fighting use is secured by tank storage. Water piping and fire hydrants are always kept at a water pressure of 7 kg/cm^2G. Fire pumps, both motor driven and diesel driven types, are kept at stand-by for immediate turnout. In addition, various types of fire trucks are stationed in the refinery.

6.2.7 Flare Stack and Blowdown System

When process units are in emergency, the gases and liquids in the units are discharged through safety valves, pressure control valves, or through emergency blowdown valves of furnaces, and are burnt at the flare stack.

6.2.8 Control Rooms

In order to integrally control the whole operations of process units, utility and offsite facilities of the refinery, the control rooms are equipped with instrumentation panels. The refinery will have three control rooms for processes, utilities and offsite facilities, respectively.

6.2.9 Buildings

Table 6.10, lists all the buildings constructed for the refinery, which are common to the three cases.

Table 6.10 Building

Building	No.s	Floor Area, m^2
Administration Office	1	3,000
Maintenance Shop	1	2,000
Warehouse	3	2,000
Laboratory	1	500
Engineering Office	1	1,000
Control Room	3	2,000
Power House	2	4,000
Substation	20	4,000
Fire House	1	500
Cafeteria	1	500
Clinic	1	300
Rest House	2	200
Gate House	2	100

6.2.10 Others

Additional requirements for facilities include communications equipment, both internal and external; lighting equipment; roads and fences; and collective stacks.

CHAPTER 7 UTILIZATION OF BY-PRODUCTS

This chapter outlines the methods of utilizing those heavy fractions by-produced from upgrading process at the Orinoco heavy oil upgrading refinery.

It is premised for the review of by-product utilization that any by-product of heavy residual is, under this study, used as boiler fuel for the purposes of steam supply to the refinery and power supply to the refinery as well as to the oil field for Orinoco crude oil production.

Information in this chapter is given in details in Sections 1.3, 2.3 and 3.3 of Volume II. In this chapter, the boiler plant capacity is adjusted in consideration of the steam balance for the entire refinery.

Basic considerations for combustion methods are given in Chapter 7, "Combustion of Heavy By-Product" of Volume II.

7.1 BY-PRODUCTS (HEAVY RESIDUALS)

Table 7.1 gives the quantities and properties of heavy residuals by-produced from the three cases of upgrading processes.

7.2 TYPES OF COMBUSTION

7.2.1 Fluid Coke

There are many experiences in boilers using fluid coke as their fuel. There is no problem in the use of boilers of the coke burning type. Fluid coke contains quite small amounts of volatile matter, and for this reason, it is less ignitable. When it is burnt, flames rise high. Ignition temperature is in the range of 870-920°C. Since the composition of fluid coke comparatively resembles that of anthracite, it is considered most appropriate to use the vertical U-type burning of boilers. This type is used for many anthracite-burning boilers. Coke has to be milled for better ignitability. About 90% of coke is required to have a particle size of 200 mesh or below. Another requirement is that auxiliary fuel gas is always burnt in an amount corresponding to 10% of the total boiler heat requirement, to make combustion stable against load fluctuations.

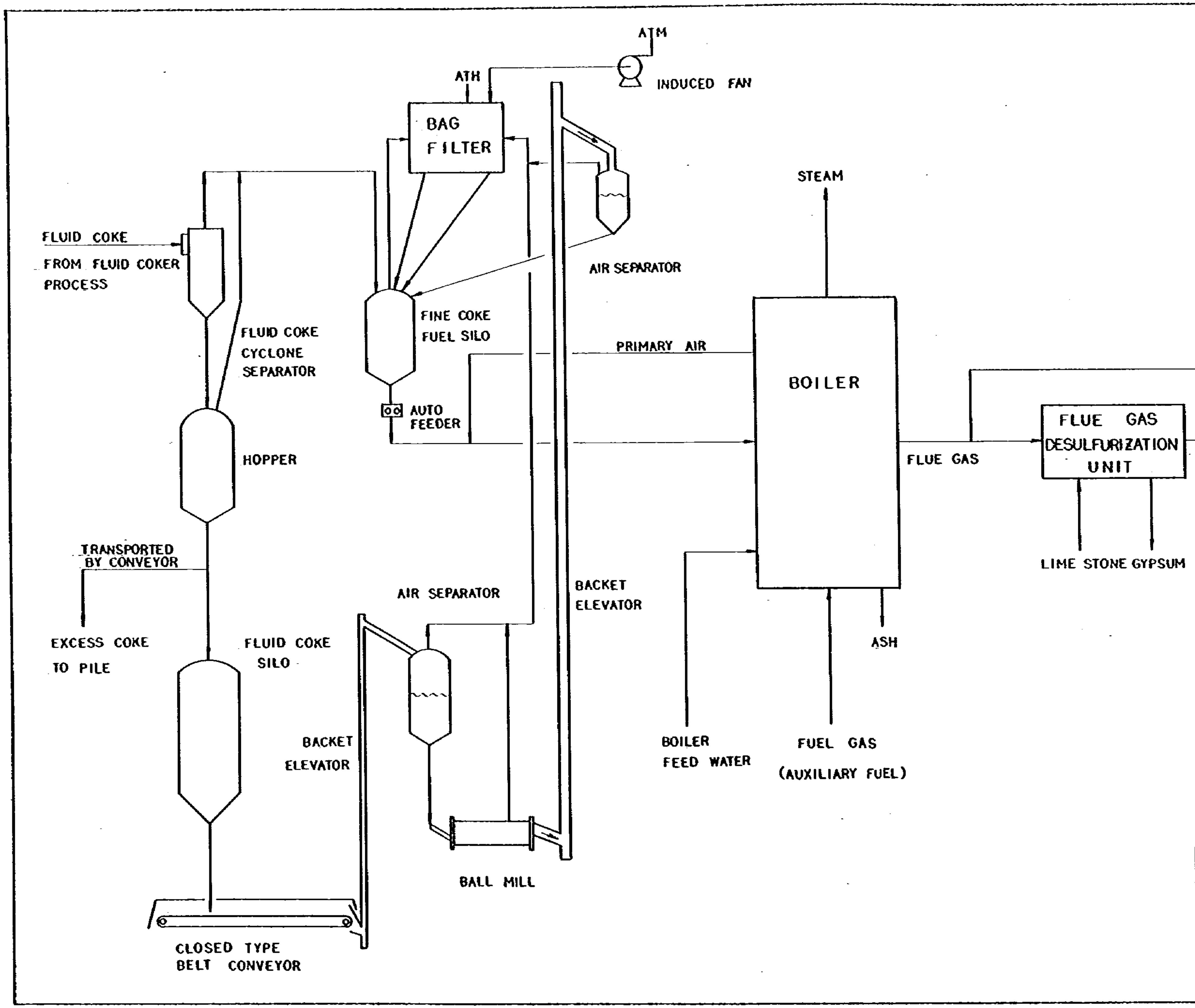
The foregoing type of combustion is achieved by the planned combustion system flow diagram shown in Fig. 7.1. Various systems may be proposed to transfer, mill and feed coke to the boiler, but there is no large difference among them.

Fluid coke withdrawn from the elutriator of the fluid coker process is air-conveyed to cyclone separators to remove fine particles. The coke of larger particle sizes is passed through hoppers into a silo. The volume of coke stored in the silo is determined by the operating conditions for both of fluid coker process and boiler plant.

Meanwhile, the surplus amount, i.e., the difference between fluid coke production

Table 7.1 Summary of By-Products

	Case	Fluid Coker	Eureka	M-DS
1.	Kind of By-product	Fluid Coke	Eureka pitch	M-DS Asphalt
2.	Total Quantity of By-Product	3,913 T/SD x 295 D/Y 1,956.5 T/SD x 70 D/Y	5,269 T/SD x 295 D/Y 2,634.5 T/SD x 70 D/Y	4,763 T/SD x 295 D/Y 2,381.5 T/SD x 70 D/Y
	Total	1,291,290 T/Y	1,738,770 T/Y	1,571,790 T/Y
3.	Balance of By-product Consumption for Boiler Fuel	1,493 T/SD x 365 D/Y	2,390 T/SD x 295 D/Y 1,910.4 T/SD x 70 D/Y	2,227.2 T/SD x 295 D/Y 1,752 T/SD x 70 D/Y
	Sub Total	544,945 T/Y	838,778 T/Y	779,664 T/Y
	Pile for stock	2,420 T/SD x 295 D/Y 463.5 T/SD x 70 D/Y	2,879 T/SD x 295 D/Y 724.1 T/SD x 70 D/Y	2,535.8 T/SD x 295 D/Y 629.5 T/SD x 70 D/Y
	Sub Total	746,345 T/Y	899,992 T/Y	792,126 T/Y
	Total	1,291,290 T/Y	1,738,770 T/Y	1,571,790 T/Y
4.	Properties of By-products	**Sulfur content** 5.79 wt.% **Metal content** V = 2,460 ppm Ni = 610 ppm Fe = 70 ppm **Bulk density** 56 LP/ft^3 (0.897 g/cm^3) **Mesh Size** Normal % on 20 (841μ) 5 50 (297μ) 15 60 (250μ) 25 80 (177μ) 55 100 (149μ) 65 140 (105μ) 75 200 (74μ) 95 Fine case % on 1,000 μ 3.2 590 μ 0.2 297 μ 0.4 250 μ 1.3 177 μ 5.5 149 μ 15.6 125 μ 24.9 105 μ 20.7 88 μ 12.5 74 μ 5.0 63 μ 6.7 53 μ 2.9 53 μ- 1.1	**Sulfur content** 4.3 wt.% **Softening point** 428°F **Volatile matter** 45.3 wt.% C 86.1 wt.% H 6.1 " S 4.4 " N 1.7 " H/C 0.85 – **Metal Content** V = 1,598 ppm Ni = 400 ppm **Heptane Insol** 72.1 wt.% **Benzen Insol.** 49.5 wt.% **Quinoline Insol.** 10.5 wt.%	**Sulfur content** 5.82 wt.% V = 1,683 ppm Ni = 393 ppm **°API gravity** –10.6 (Sp. Gr. = 1.1706) R & B Soft. Pt. 162°C Nitrogen 1.58 wt.% Asphalten (C_7) 43.3 wt.% Con. Carbon 57 wt.% Viscosity 4,000 cp @ 250°C 500 cp @ 300°C



ORINOCO HEAVY OIL UPGRADING PROJECT

SIMPLIFIED FLOW OF COMBUSTION SYSTEM FLUID COKER CASE

FIG. 7.1

and boiler fuel requirement, is not used for the time being, and therefore, is brought from the hoppers by means of a belt conveyor to the outside of the refinery for outdoor storage.

The coke stored in the silo is sent by means of a belt conveyor and a backet elevator to an air separator to remove fine particles. The coke of larger particle sizes is then ground in ball mills. Coke is again passed through another air separator, and fine coke ready for use as boiler fuel is stored in fine coke silo.

Coke is fed to the boiler by an automatic feeder which air-conveys the coke feed, using boiler primary air. The boiler is additionally supplied with off-gas from the refinery, as an auxiliary fuel.

Flue gas coming out of the boiler is treated by the flue gas desulfurization unit and discharged to the atmosphere through the stack.

The following table gives steam and power requirements and the balance between production and consumption of by-product coke, both in normal operation (100%) and during periodical maintenance (50%).

	100% operation (295 days/yr.)	50% operation (70 days/yr.)
Steam requirements:		
100 kg/cm^2G	772.1 T/H	772.1 T/H
50 kg/cm^2G	388.4 T/H	194.2 T/H
Power requirements:		
Total	200,400 KW	182,700 KW
(Oil field)	(126,200 KW)	(126,200 KW)
Coke production	3,913 T/SD	1,956.5 T/SD
Coke quantity for boiler fuel	1,493 T/SD	1,493 T/SD
Surplus coke	2,420 T/SD	463.5 T/SD
Volume of boiler fuel gas	0.11 $MMNm^3/SD$	0.11 $MMNm^3/SD$
Volume of CO boiler fuel gas	0.33 $MMNm^3/SD$	0.165 $MMNm^3/SD$
Surplus volume of fuel gas	1.08 $MMNm^3/SD$	0.485 $MMNm^3/SD$

Taking the above requirements and balance into consideration, this study plans to generate the steam of 100 kg/cm^2G and 500°C from boilers using coke fuel and to generate the steam of 50 kg/cm^2G. and 405°C from CO boilers using CO gas derived from coke combustion. Major machinery in the combustion system of the fluid coker process is as follows:

Fluid coke cyclone separators

Hoppers

Belt conveyor for surplus coke
Fluid coke silos
Closed type belt conveyors
Ball mills
Backet elevator
Air separator
Induced fans
Bag filters
Fine coke silo
Boilers
(The flue gas desulfurization unit is dealt with as an offsite facility.)

7.2.2 Eureka Pitch

Eureka pitch has a larger calorific value and gives less ash than does coal. It is said to be pulverized. Pitch is in the solid form at normal temperature, but it can be handled in the liquid form above its softening point. It is possible, therefore, to use the following two pitch-burning methods:

– Atomized liquid pitch burning.
– Pulverized solid pitch burning.

The latter type of burning has been adopted under this study for the convenience of pitch transport and storage.

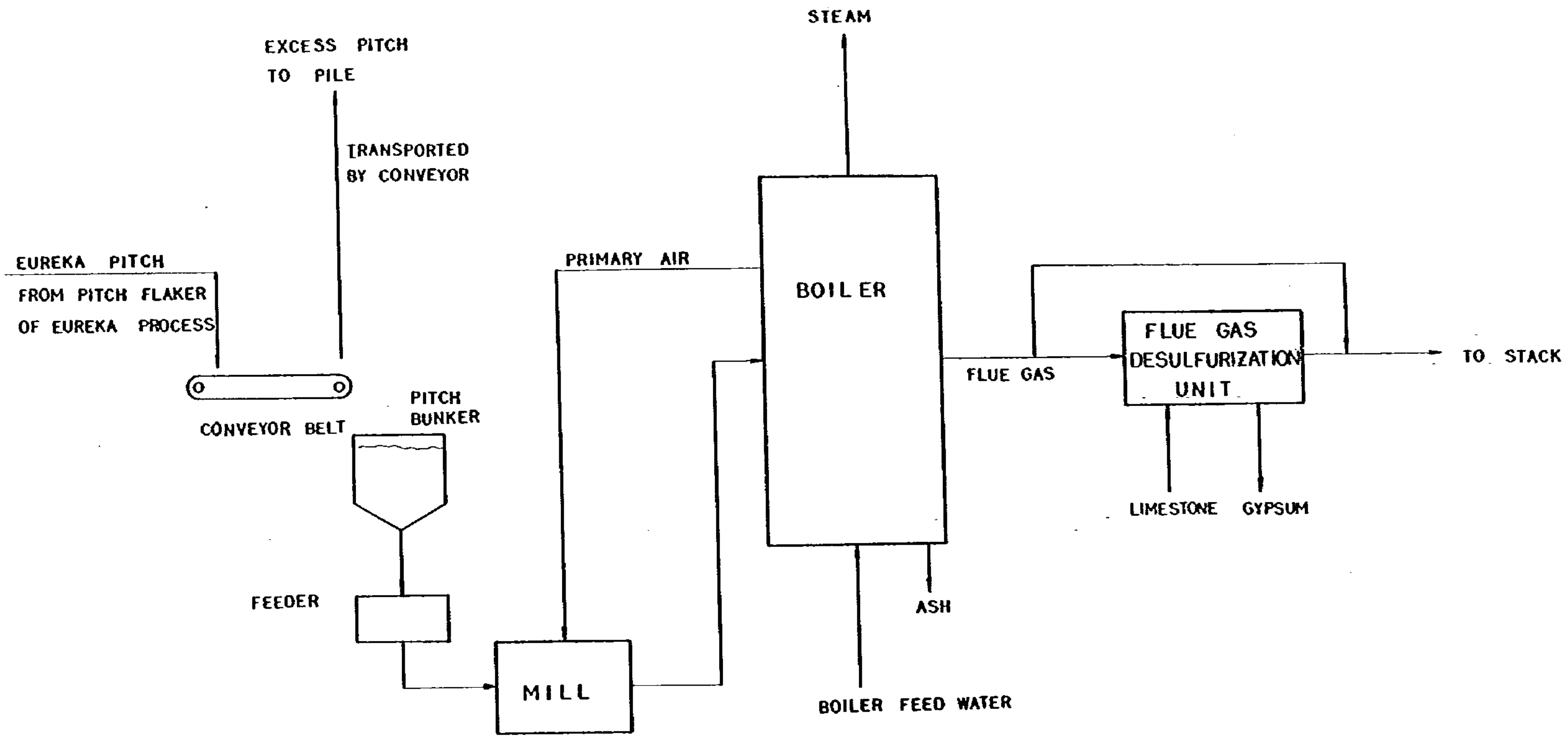
Like fluid coke, pitch will pose no problem as long as pitch is pulverized as in the fine coke burning method. Eureka pitch burning is considered to required no auxiliary fuel.

The foregoing type of combustion is achieved by the planned combustion system flow diagram shown in Fig. 7.2

The pitch by-produced in the Eureka process comes out of the pitch flaker. It is transported to a pitch bunker by way of a belt conveyor. The surplus amount other than boiler fuel requirement is directly belt-conveyored to an outdoor storage yard outside the refinery, where pitch is stock-piled for future use.

Pitch has a high softening point, and is less susceptible to natural oxidation than coal. Since it has little fear of heat accumulation and natural ignition, it can be stored in an open yard. Pitch is passed from the pitch bunker to the feeder, where pitch is weighed and removed of impurities. The feeder also automatically controls the volume of pitch fed to the boiler.

Pitch leaves the feeder and goes to the mill, where pitch is pulverized and simultaneously dried by hot primary air from the boiler. This fine pitch is then fed to the boiler.



ORINOCO HEAVY OIL
UPGRADING PROJECT

SIMPLIFIED FLOW OF
COMBUSTION SYSTEM
EUREKA CASE

FIG. 7.2

The following table gives steam and power requirements and the balance between production and consumption of the by-product pitch, both in normal operation and during periodical maintenance, for the refinery.

	100% operation (295 days/yr.)	50% operation (70 days/yr.)
Steam requirements:		
100 kg/cm²G	1,190.3 T/H	951.4 T/H
50 kg/cm²G	0 T/H	0 T/H
Power requirements:		
Total	208,000 KW	182,500 KW
(Oil field)	(126,600 KW)	(126,600 KW)
Pitch production	5,269 T/SD	2,634.5 T/SD
Pitch quantity for boiler fuel	2,390 T/SD	1,910.4 T/SD
Surplus pitch	2,879 T/SD	724.1 T/SD

Major equipment in the pitch combustion system is as follows:

Conveyor belt

Pitch bunker

Feeder

Mill

Boiler

(The flue gas desulfurization unit is dealt with as an offsite facility)

7.2.3 M-DS Asphalt

As regards the combustion of M-DS asphalt, Maruzen Oil has reviewed the following four methods:

Fluidized-bed combustion

Pulverized solids combustion

Low-viscosity oil cutback, and

High-temperature atomization.

It has been revealed that:

- The fluidized-bed combustion method is at present no established technology, although it is promising in the future.
- The pulverized solids combustion method, composed of almost existing technologies, is highly practicable, but it calls for large facilities for solids handling in such steps as flake formation from molten asphalt, pulverization of flake asphalt, and storage of powdery asphalt in silos.
- The low-viscosity oil cutback method has a merit of permitting the use of ordinary atomizing methods, but vacuum gas oil, i.e., the valuable base oil of improved crude

oil, is consumed as fuel.

In the light of these review results, Maruzen Oil has now studied a method of burning asphalt by atomizing the liquid form of asphalt, in order to take full advantage of the solvent deasphalting process.

A small test combustion furnace was used to carry out atomizing/burning burner tests with the asphalt having a softening point of about 160°C obtained from the deasphalting process. It has been demonstrated that the same combustion as in the use of ordinary fuel oil can be accomplished by heating asphalt to decrease its viscocity to a suitable level for atomizing and by selecting those burners that enable flames to be fully swirled. Maruzen Oil thus proposes the following combustion system.

(1) Asphalt is withdrawn from the asphalt stripper through the bottom at a high temperature of about 340°C, and it is directly transported to the burners through the pipe jacketed with hot oil, without being passed through an intermediate tank and a service tank.

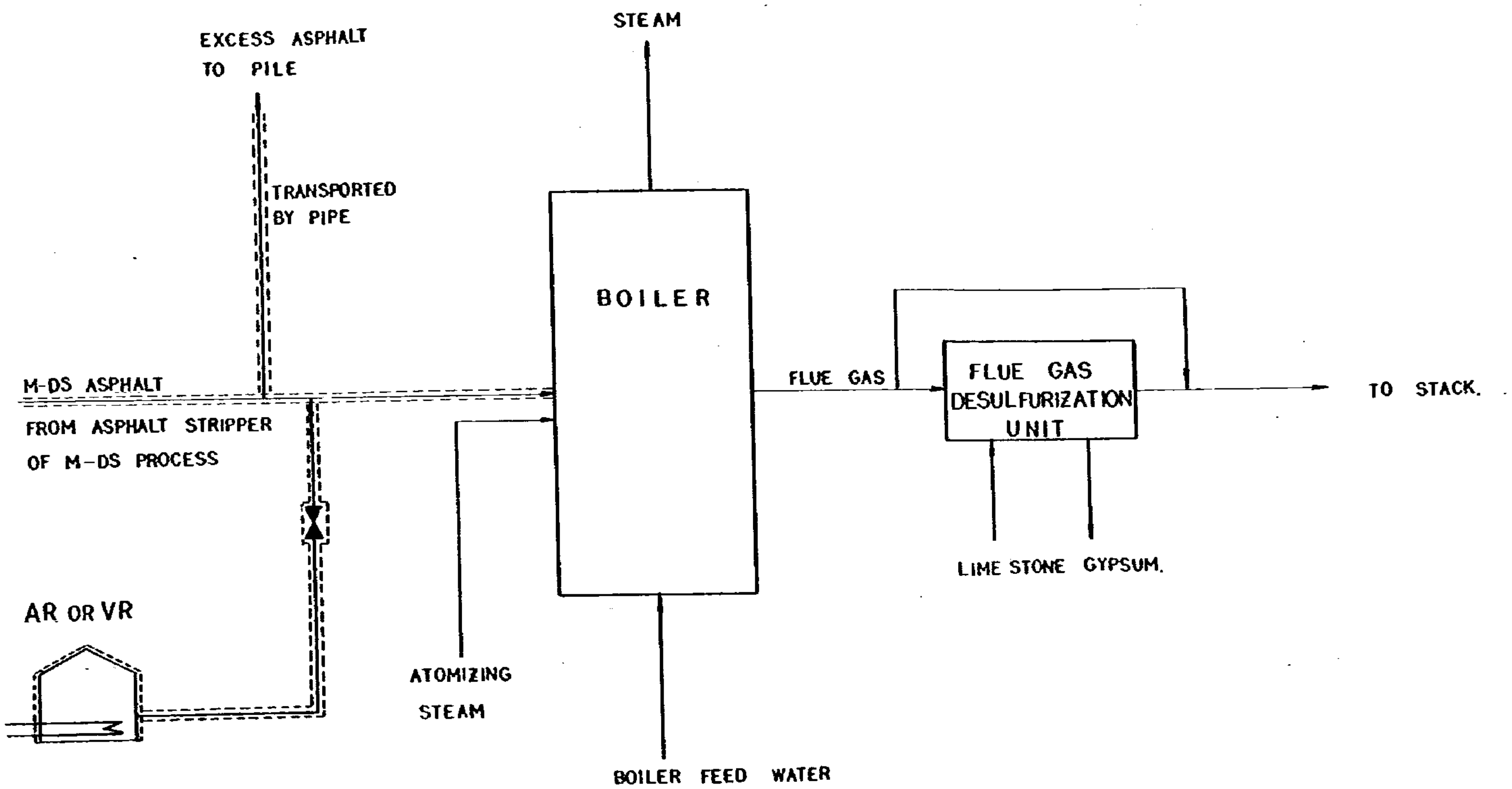
(2) High-temperature steam is used to heat asphalt fuel and atomize it through internal mixing burners and to maintain an atomizing temperature.

The direct link between the deasphalting unit and the boiler may allow the boiler operation to be directly affected by the operation of the deasphalting unit. Therefore, the asphalt combustion system is also provided with the piping that enables vacuum residue to be used as an alternative fuel in such a case as emergency shutdown of the deasphalting unit.

For this reason, a vacuum residue tank is installed as an intermediate tank, instead of installing an intermediate tank of asphalt. When deasphalting unit is not in normal operation, such as start up or shut down, the asphalt produced is not fed directly to the boiler, but is mixed with deasphalted oil and the mixture is sent to the vacuum residue tank. Fig. 7.3 shows the flow diagram for the asphalt combustion system.

The following table gives steam and power requirements and the balance between production and consumption of the by-product asphalt, both in normal operation (100%) and during the time of 50% operation.

	100% operation (295 days/yr.)	50% operation (70 days/yr.)
Steam requirements:		
100 kg/cm²G	1,180.4 T/H	928.5 T/H
50 kg/cm²G	0 T/H	0 T/H
Power requirements:		
Total	203,000 KW	176,000 KW
(Oil field)	(120,500 KW)	(120,500 KW)
M-DS asphalt production	4,763 T/SD	2,381.5 T/SD
M-DS asphalt quantity for boiler fuel	2,227.2 T/SD	1,752 T/SD
Surplus asphalt	2,535.8 T/SD	629.5 T/SD



ORINOCO HEAVY OIL UPGRADING PROJECT

SIMPLIFIED FLOW OF COMBUSTION SYSTEM M-DS CASE

FIG. 7.3

CHAPTER 8 PROJECT EXECUTION

A preparative study on the project execution is described in this chapter, with a view to setting up criteria for various cost estimation to be described in Chapters 9 and 10. This chapter therefore deals with the descriptions on the following items:

- Refinery plot plans
- Refinery construction schedule
- Mobilization plans
- Refinery organization
- Employee training plan

The refinery under this study is a special refinery characterized by upgrading heavy crude oil to obtain improve crude oil. What is more, the refinery stands on the pre-requisite condition that it is going to supply the oil field with diluent, power, and fuel gas necessary for crude oil production. Therefore, execution of this project inevitably calls for a comprehensive study in which both the oil field and the refinery are taken into due consideration. The refinery to be constructed will have to have a large scale, if consideration is also given to the special process for upgrading heavy oil and to power generation by means of special boilers using a process by-product as the fuel.

8.1 REFINERY PLOT PLAN

Fig. 8.1 shows the fluid coker case of plot plan for the entire refinery. Refinery site is required to accommodate raw material storage facilities, process units, intermediate and product tanks, storage facilities other than tanks, utility facilities, and offsite facilities.

Land requirements for the three cases are almost same and about 1,500,000 m^2. In the case of this refinery, the major product of improved crude oil is almost immediately pipelined to the shipping port. This permits the refinery to have quite a smaller tank capacity within the refinery, as compared with those refineries from which the products are directly pipelined to tankers for loading. The same applies to the storage capacities of other by-products.

Sulfur, gypsum and fuel by-products may require a further detailed plot plan depending on the solids storage conditions.

8.2 REFINERY CONSTRUCTION SCHEDULE

An overall construction schedule must be prepared on the basis of present situations in, and planned target date of completion for, the refinery construction project, as well as on the basis of individual review by project participants, including technical development and basic design conducted by process licensors, and the design, procurement, transportation, construction and operation conducted by the contractor. The schedule described in this Section is a tentative construction schedule, common to all the three cases, prepared under the following assumptions:

- For all the three cases, the refinery is different only in the major upgrading process units, and there is no large difference from a construction scheduling point of view.
- The planned target date of mechanical completion is set for late 1987.
- The contractor is awarded the contract for refinery construction at such a time as required to construct it within the shortest period of construction work and to complete the work successfully by the target date.
- Process unit licensors are selected, and basic design data are available, before the contractor has been awarded with the contract.
- Basic design data and general specifications of the project facilities are already established at the time of contract awarding.
- Orders for major refinery equipment is put as early as possible, because the delivery of such equipment constitutes a bottleneck in the construction schedule. And the equipments having many unit numbers are to be ordered separately to some manufactuers.
- Orders for major refinery equipment are awarded by calling for bids on such equipment during a suitable period.
- Equipment and machinery shall be delivered within average periods so far experienced with similar types.
- Infrastructures, such as land prepared in good conditions, roads extending to the construction site, etc., are ready for use by the time they become necessary.
- Refinery is supplied with crude oil concurrently with the completion of construction work, but the operating rate is set at 50% for the first year on the premises that the refinery cannot be put into full operation during the initial half year.

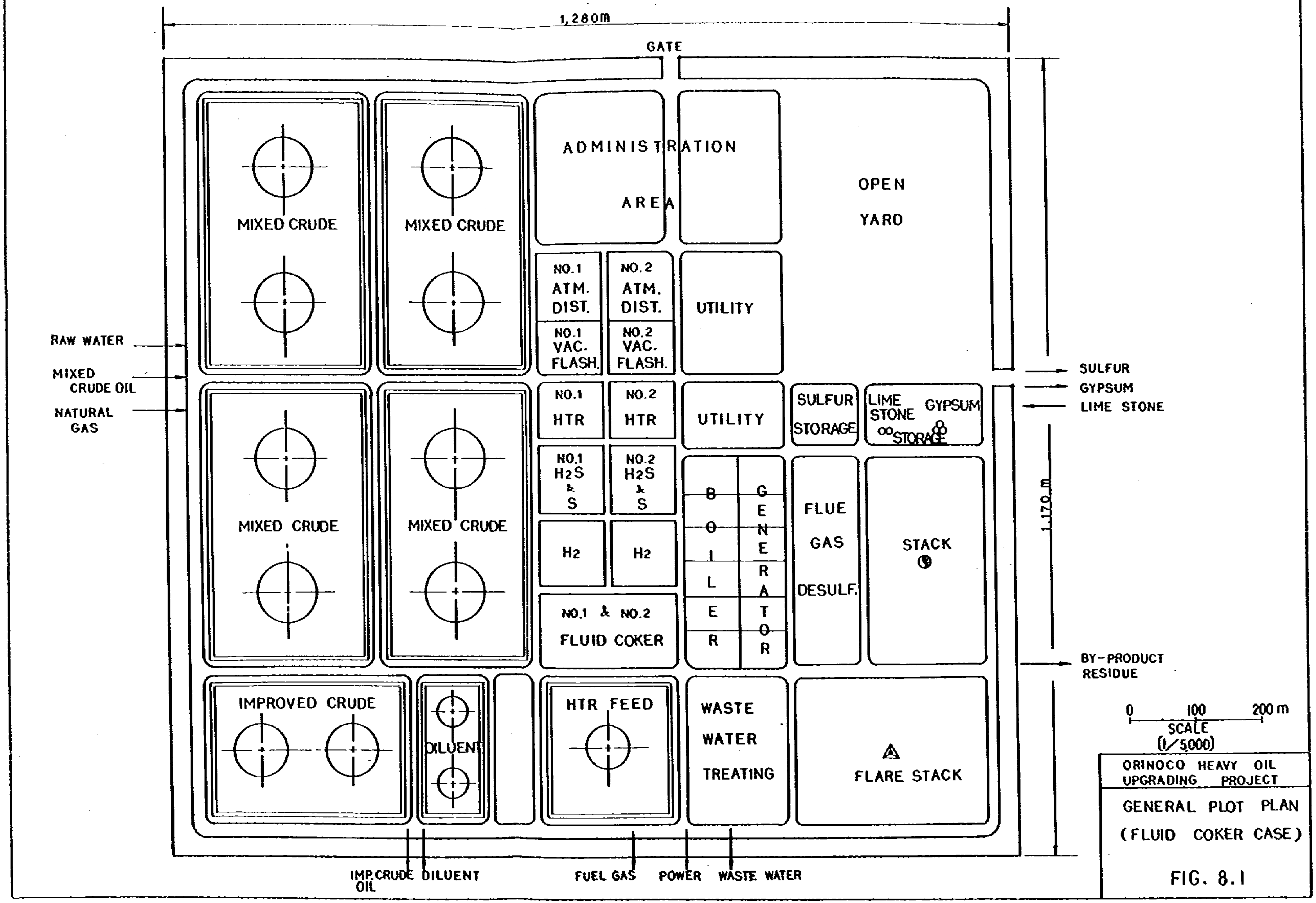
Fig. 8.2 shows the entire refinery construction schedule prepared on the basis of the baove assumptions.

8.3 MOBILIZATION PLAN

In carrying out construction work effectively for the upgrading refinery, it is necessary to estimate how many workers can be put into construction work throughout the construction period, as well as during the peak period(s). Estimated number of workers offers an aid to the plans on accompanying infrastructures. Studies will be made also on the procurement of construction equipment, as well as on unloading and land transportation of equipment and materials.

8.3.1 Labor Mobilization

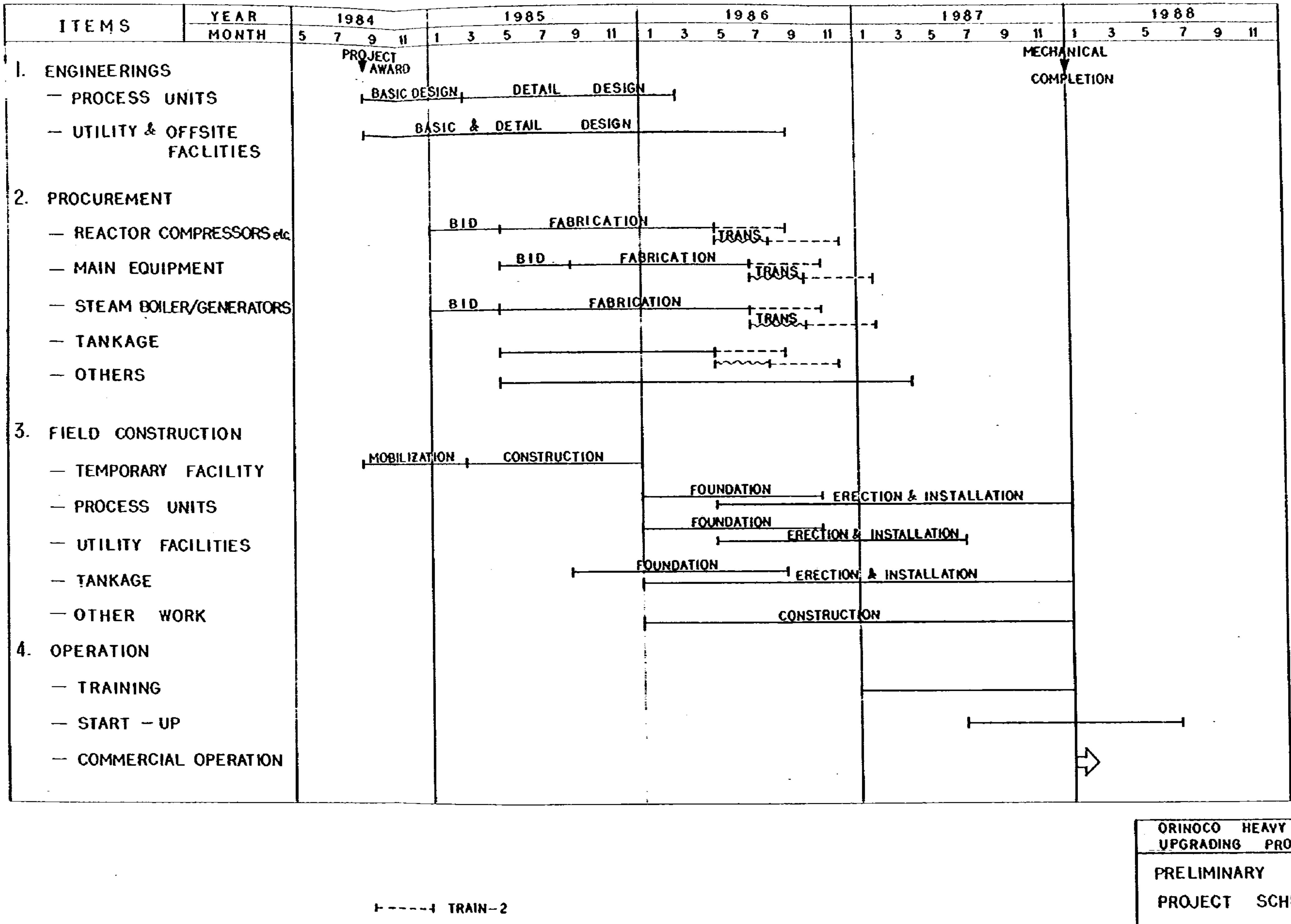
Workers engaged in the refinery construction include those supervisors who supervise and coordinate the entire construction work as the work is under way at the



ORINOCO HEAVY OIL UPGRADING PROJECT

GENERAL PLOT PLAN (FLUID COKER CASE)

FIG. 8.1



ORINOCO HEAVY OIL UPGRADING PROJECT

PRELIMINARY PROJECT SCHEDULE

FIG. 8.2

construction site, along with those workers who are directly engaged in the work, such as foremen, skilled labor, semi-skilled labor, and unskilled labor.

As regards the mobilization situations, a peak will come during a period from late 1986 to early 1987, and 3,500 to 4,000 workers will be required during that period. The labor requirement throughout the construction period is estimated at about a total of 2.0 million man-days.

8.3.2 Procurement of Construction Equipment

An on-the-spot survey has revealed that in Venezuela, the costs of construction equipment on lease are often calculated on the basis of 10 months' depreciation period. Naturally, the lease is very costly. In large-scale construction work such as contemplated under this project, it seems rather reasonable to buy new equipment in Venezuela or import new equipment from abroad.

As for the equipment required to install refineary units on the Orinoco construction site, the sizes of such equipment must be in good match with the sizes and weights of the units.

The maximum size of equipment is determined by restrictive conditions for inland transportation. If a equipment is larger than the limit set by transportation, it is divided into several pieces until they are brought into the site. They are then reassembled or reconnected for installation on the site. It will thus become necessary to use such construction equipment as adequate for this knock-down method of installation.

This precaution is not required if the roads for inland transportation to the refinery is newly constructed, including bridges and/or overbridges, if any.

8.3.3 Unloading and Inland Transportation of Equipment and Materials

Equipment and materials imported under this project will be unloaded and cleared at Pto. Ordaz or Pto. La Cruz area.

A wharf with unloading facilities will be constructed and used exclusively for the unloading operations under this project. Therefore, unloading capacity will pose no problem.

As regards the inland transportation from the unloading wharf to the Orinoco construction site, it is estimated from an on-the-spot survey that, except for the case of new road construction, the maximum size of transportable equipment will be 3,000 m/m in diameter and 30,000 m/m long, with a weight of 80 tons per shipment.

8.4 REFINERY ORGANIZATION

This section deals with an organization and a manning plan required to operate the refinery. This organization and associated manning plan are worked out on an assumption that the refinery under study is one of the refineries operated by the Venezuelan company concerned and therefore that general administration of the refinery under study will be integrally controlled at its headoffice. Even if personnel engaged in general administration is placed actually in the refinery, this study does not include such personnel in this organization, nor include labor cost for such personnel in the operating cost.

As detailed in Fig. 8.3, the refinery organization largely consists of the three departments of Operation Department directly in charge of operation, Technical Department, and Maintenance Department.

The refinery organization is headed by a refinery manager, to whom three department managers directly report. These three department managers are responsible for the activities of their respective departments.

Table 8.1 shows the number of personnel required to operate the refiery under this organization. There is little difference of personnel requirement among the three upgrading process cases. Thus the number in this table can be regarded as common to all cases.

Technical Department:

Technical Department is organized by three sections of Technical service, Production Planning and Laboratory. The Department is in charge of the following tasks:

- Setting of various operating standards.
- Planning on process unit operations.
- Preparation of production plans.
- Planning, review, and execution of maintenance, improvement and trouble-shooting for the refining.
- Technical studies.
- Operation and administration of laboratory

Technical Department covers all the technical matters associated with refinery operation, as well as with product quality control.

Operation Department:

Operation Department, organized by Process Section, Utility Section, and Offsite Section, is in charge of the following tasks:

- Operation and maintenance of process units.
- Operation and maintenance of utility facilities.
- Operation and maintenance of offsite facilities, including oil blending and transport, receiving of crude oil and shipment of products, and storage and shipping of other materials.
- Preparation of oil handling plans.
- Preparation of statistics recording refinery operation, material receiving and shipment, etc.

Table 8.1 Summary of Required Personnel

	Personnel Class	Manager Class	Supervisor Class	Foreman Class	Operator Class		Clerk & Worker Class
	Operation	Day	Day	Shift	Shift	Day	Day
1.	Refinery Manager	1					
	Assistant Manager	1					
	Secretary					2	
2.	Technical Department	2					2
	Technical Service Sect.		12			4	2
	Production Planning Sect.		4				1
	Laboratory Sect.		2		4 X 5	4	1
3.	Operation Department	2					2
	No. 1 Process Sect.		2	1 X 5	10 X 5		1
	No. 2 Process Sect.		2	1 X 5	10 X 5		1
	No. 3 Process Sect.		2	1 X 5	6 X 5		1
	Boiler/Generator Sect.		2	1 X 5	6 X 5		1
	Utility Sect.		2	1 X 5	6 X 5		1
	Offsite Sect.		2	1 X 5	6 X 5	20	1 + 30
4.	Maintenance Department	2					2
	Maintenance Sect.		2	1 X 5	2 X 5	45	3 + 15
	Mechanical Sect.		2			21	2 + 10
	Instrument Sect.		2	1 X 5	2 X 5	10	1
	Electrical Sect.		2			16	1
	Warehouse Sect.		2			4	1 + 8
	Total	8	40	40	260	126	87
	Grand Total		561				

Operation Department is directly responsible for refinery operation. The Department gets prepared for continuous operation in 5 shifts in every 24 hours.

Maintenance Department:

This Department, organized by five sections of Maintenance, Mechanical, Instrument, Electrical, and Warehouse, is in charge of the following tasks:

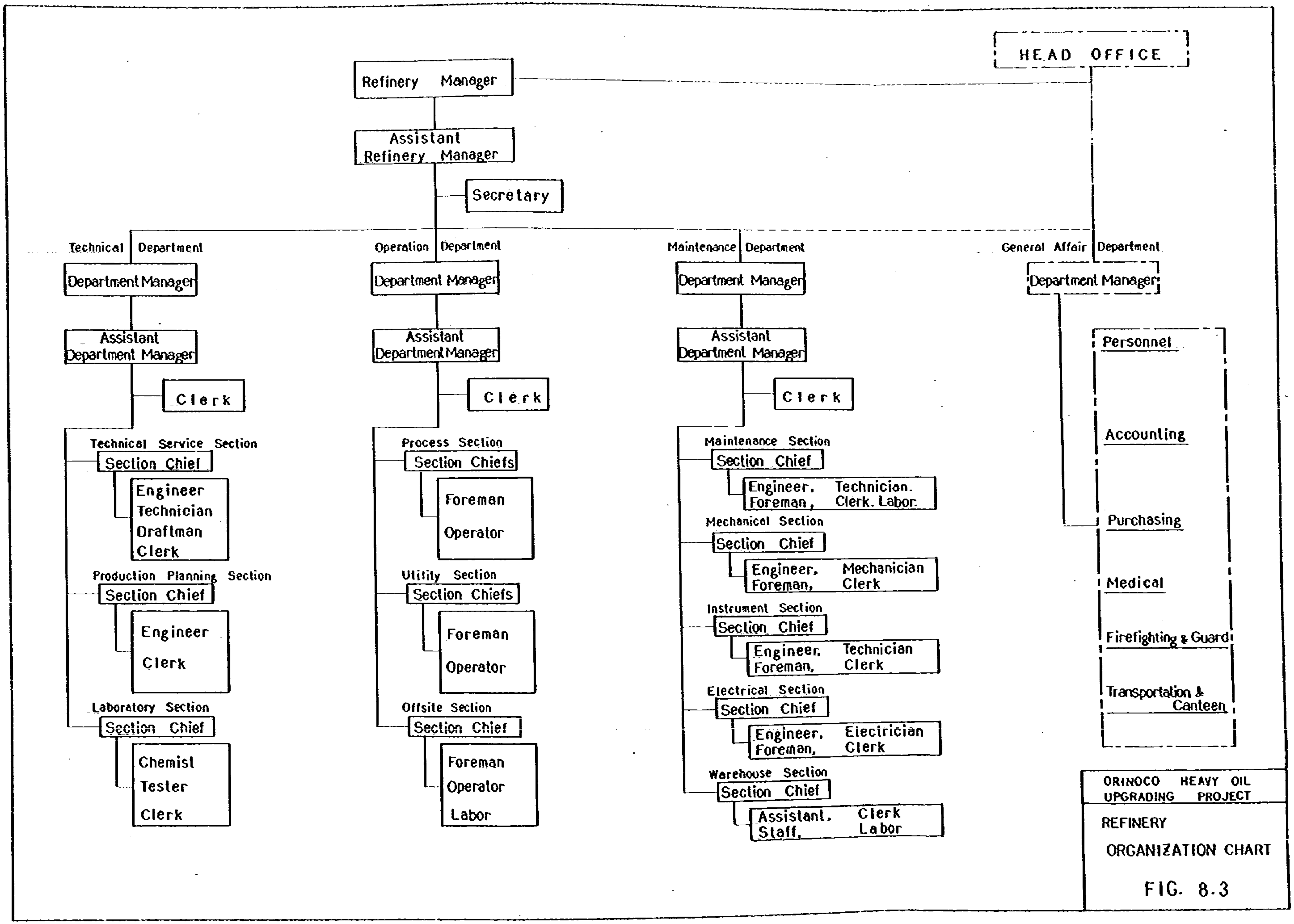
- Planning on maintenance and repairs of equipment.
- Preparation of execution plans.
- Routine repair work.
- Ordering for minatenance work to outsiders, supervision and inspection of such work.
- Preparation of statistics on maintenance work.
- Setting of standards associated with maintenance.
- Preparation and control of parts- and material-purchasing plans.
- Preparation of statistics on maintenance materials.

Maintenance Department gets prepared for accomplishing repair work, deassembly and repairs of machinery, other than large work conducted at the time of periodical maintenance. A part of workers in this Department need working on a shift basis for the repair work during operation.

General Affairs Deaprtment:

Although this Department is not included in the refinery organization, its roles, such as listed below, must be accomplished for smooth operation of both the refinery and the owner company. If the headoffice of the company regards a refinery as a division, the General Affairs Department is often incorporated in the refinery organization. In this study, however, this department is excluded from the refinery organization, as intended by the Venezuelan side. The department is in charge of the following tasks:

- Personnel control and salary payments.
- Employee welfare.
- Planning and execution of personnel education and training.
- Budgeting and accounting.
- Administration of land and buildings.
- Guard keeping and anti-pollution measures.
- Fire-fighting.
- Clinic operation and control; Employee safety and hygienic control.
- Public relations.
- Mess hall operation and administration.
- Canteen.
- Other general affairs.



ORINOCO HEAVY OIL UPGRADING PROJECT

REFINERY ORGANIZATION CHART

FIG. 8.3

8.5 EMPLOYEE TRAINING PLAN

Smooth operation of the refinery can be secured by training refinery employees. Thus, a training plan is required for the personnel described in the previous section. The cost of this training plan is included in the capital cost as one of the pre-startup costs.

The trainees include both the foreman class and the operator class of Operation Department. Among operators, chief operators serve as the key person of each shift group, and they need especially long-term training.

The method and period of training are based on the following program:

- The class of foremen who operate process units:

  They are trained first at a similar process plant or a model plant for 6 months, to learn refinery operation. After this period, they are allow to attend the construction work of their own refinery for another 6 months, during which period they learn the organization and functions of the new refinery.

- The class of foremen who operate utility and offsite facilities:

  These foremen attend the construction work of their refinery for 6 months to learn the organization and functions of utility and offsite facilities for which they are responsible.

- Chief operator class of Operation Department:

  Chief operators in this class attend basic lectures on the operation of the refinery, and then they are given more advanced lectures on their respective expertise. After lectures are complete, they are given actual field training at an existing refinery in Venezuela. It will take 12 months to complete this course.

- General operator class:

  Operators are trained at the construction site for 6 months.

- Laboratory testers of Technical Department:

  Testers are lectured and field-trained at an existing refinery in Venezuela for 6 months.

As regards the management class and the employees with whom Technical and Maintenance Departments are manned, it is assumed that they already have sufficient technical knowledge. Therefore, they are given no special training.

The foregoing program should be carried out prior to the startup of refinery. After the refinery has been put into operation, all employees are trained with field practice during 6 months of test operation period. Table 8.2 gives the number of trainees and the periods of training for the respective classes.

Table 8.2 Trainees

Class of Trainee		Number of Trainee	Training Months	Training Place
1.	Foreman of Process Sect.	15	6	Existing Refinery
			6	Plant Site
2.	Foreman of Utility & Offsite Sect.	15	6	Plant Site
3.	Chief Operator of Operating Department			
	process sect.	25	12	Existing Refinery
	utility sect.	10	12	Existing Refinery
	offsite sect.	5	12	Existing Refinery
4.	Operator of Operating Department			
	process sect.	105	6	Plant Site
	utility sect.	50	6	Plant Site
	offsite sect.	25	6	Plant Site
5.	Tester of Laboratory sect. in Technical Department	24	6	Existing Refinery
	Total	274	–	–

CHAPTER 9 CAPITAL REQUIREMENTS

This chapter deals with the capital requirements for the construction of Orinoco heavy oil upgrading refinery which has been outlined up to Chapter 8, above. The capital requirement refers to a sum of capital to be invested by the time when the refinery is put into commercial operation. It is largely divided into fixed capital (subject to depreciation) and working capital (not subject to depreciation). These two types of capital include the following items:

Fixed capital:

- Construction cost
- Paid-up royalties
- Initial costs of catalyst and chemicals
- Pre-operating expenses

  (As regards the interest rates imposed during construction period, the total capital requirement is covered by fund on hand, and therefore, interest is not included in the fixed capital.)

Working capital:

- Land
- Raw material and product inventories.
- Catalyst and chemical inventories.
- Spare parts and warehouse supplies.
- Cash in hand.
- Balance of accounts receivable and accounts payable.

Table 9.1 gives a summary of capital requirements.

Table 9.1 Capaital Requirement Summary

		Fluid Coker Case	Eureka Case	M-DS Case
		(10^6 US$)	(10^6 US$)	(10^6 US$)
1.	Process Unit	460.6	452.2	554.7
2.	Utility Facilities	274.5	320.0	281.3
3.	Offsite Facilities	187.1	177.8	173.8
4.	Paid-up Royalties	2.09	2.13	2.39
5.	Initial Catalyst and Chemicals	5.73	5.10	25.26
6.	Pre-operating Expenses	16.40	16.64	16.64
	Fixed Capital (1 – 6)	946.42	973.87	1,054.09
7.	Working Capital	126.98	123.63	134.09
Total Capital Requirements		1,073.40	1,097.50	1,188.18

The subsequent Sections 9.1 and 9.2 will describe the standards and methods used to calculate respective cost items, and will give estimated costs.

9.1 Fixed Capital

9.1.1 Construction costs

Construction costs have been estimated on the following premises:

(1) Construction costs are estimated on an assumption that the plant concerned is constructed in the Orinoco area.

(2) Construction costs are calculated on the "present price" basis, wherein equipment/material costs and labor cost, as of middle 1980, are used. Future changes in prices are not taken into account.

(3) Construction costs include construction material cost, labor cost, design & engineering fee, and contractor expenses.

(4) The construction material/labor costs include total direct material/labor costs, wherein battery limit units, such as listed below, are installed, and also include include indirect field expenses and worker wages.

The battery limit units involved are as follows:

Furnaces	Fire protection work
Tower, reactors and internals	Paving and concrete works
Heat exchangers.	Compressor shelters
Pumps	Control houses
Motors	Catalyst handling equipment
Compressors	Construction equipment
Pipings	Machinery
Instruments	Temporaly offices, warehouses and locker rooms
Electrical devices	Field tests
Insulation work	Tools
Structural steel work Final washing	Work-related expenses
Othe field expenses Special wages	Others

These items apply similarly to utility and offsite facilities.

(5) Design & engineering costs and contractor expenses have been estimated from past experiences. These must be added to obtain an estimated total construction cost. They include the following items:

- If process owners are involved, their costs for basic design and specifications, and their cost for check on the contractor's detailed design.
- The contractor's costs for detailed design, procurement, expediting, inspection, construction machinery and tools, office expenses, construction work supervision, and indirect contractor expenses.

(6) The following items are not included in the estimated construction cost, assuming that they can be used as established infrastructures:

- Land preparation for construction work.
- Geological surveys.
- Special foundation work.
- Port facilities.
- Roads extending up to the refinery.
- Head offices.
- Formalities, taxes and expenses inherent to the region.

(7) Procurements are based on bids called by the contractor.

(8) There would be no delay in construction work caused by weather conditions.

(9) Material and labor costs are based on estimated values.

(10) This construction cost does not include the cost of facility required to be

constructed outside the battery limit (fences) of the upgrading refinery to connect the refinery with outside facilities. Table 9.2 gives detailed breakdowns of these costs.

9.1.2 Paid-up Royalties

Royalties are individually agreed upon between a licensor and a licensee. For the convenience, estimated, figures are presented here. They constitute present values in 1980. In this study, no escalation of royalties is taken into account. Table 9.3 gives estimated paid-up royalties.

9.1.3 Initial Cost of Catalyst and Chemical Filling

This is the costs of catalysts and chemicals to be filled prior to the operation of refinery. These costs are based on present prices (1980) in Venezuela. No escalation is taken into account. Table 9.4 summarizes estimated costs.

9.1.4 Pre-operating Expenses

Pre-operating expenses are required during construction period until test runs are finished. They include the following items:

(1) Operator training cost:

This cost comprises salaries paid to the trainees during their training period and the training expenses. The cost is calculated on the basis of a personnel training plan described in Chapter 8.

(2) Administrative cost:

It is customary to account the salaries of refinery management and administrative staff and related indirect cost within the pre-operating expenses. In the case of this project, however, it is premised that these costs, required in the initial period prior to construction work and during the construction work, are regarded as an administrative cost of company head office. Thus, the administrative cost of the refinery is taken into account only for a year prior to the startup of commercial operation.

(3) Test running cost

The costs calculated under this category include the costs for those persons sent by process licensors and the contractor to help test-run the refinery, in addition to the costs of chemicals and utilities consumed during test runs.

Table 5.5 summarizes the estimated pre-operating expenses.

Table 9.2 Construction Costs Summary

	Capacity per unit in	Fluid Coker Case			Eureka Case			M-DS Case		
		Capacity	No.s	Cost 10^6 US$	Capacity	No.s	Cost 10^6 US$	Capacity	No.s	Cost 10^6 US$
1. Process Units										
Atmospheric Distillation	BPSD	102,800	2	56.5	103,200	2	56.6	98,200	2	55.0
Vacuum Flashing	BPSD	67,200	2	40.3	67,400	2	40.5	64,200	2	39.3
Fluid Coker	BPSD	43,600	2	172.1	–	–	–	–	–	–
Eureka	BPSD	–	–	–	42,400	2	151.9	–	–	–
M-DS	BPSD	–	–	–	–	–	–	40,700	2	76.1
HTR/HDS	BPSD	60,900	2	101.8	16,100	2	22.0	10,700	2	18.5
	″	–	–	–	45,900	2	88.9	48,800	2	255.8
H_2	$MMNm^3/SD$	0.90	2	51.8	1.08	2	57.9	1.70	2	76.0
Acid Gas Treating	T/SD as H_2S	67/23/192	2	19.6	315	2	14.5	309	2	14.4
Sulfur	T/SD as S	255	2	18.5	285	2	19.9	279	2	19.6
Sub-total				460.6			452.2			554.7
2. Utility Facilities										
Steam Generator System	T/H	260/200	4/2	117.2	240	6	144.7	240	6	123.0
Power Generator System	KW	55,000/ 18,000	4/2	113.0	46,000	6	125.1	44,000	6	118.5
Water Treating System	T/H	2,120	1	5.7	2,310	1	6.2	1,910	1	5.2
BFW Treating System	T/H	250	3	17.1	310	3	20.5	200	3	15.0
Condensate Recovery System	T/H	170	2	1.9	200	2	2.0	250	2	1.9
Cooling Water System	T/H	18,000	2	14.9	20,000	2	16.5	15,500	2	12.9
Fuel System	–	–	–	0.5	–	–	0.7	–	–	0.6
Air System	Nm^3/H	1,800	3	1.1	2,000	3	1.2	1,900	3	1.1
Inert Gas System	Nm^3/H	350	2	3.1	350	2	3.1	350	2	3.1
Sub-total				274.5			320.0			281.3
3. Offsite Facilities										
Storage	10^3 kl	1,436		86.2	1,397		83.4	1,340.5		80.5
Loading & Unloading	KW	6,713		3.8	5,928		1.6	57.69		1.6
Waste Water Treating	T/H	181	2	11.8	139	2	10.1	193	2	12.3
Flue Gas Desulfurization	10^6 Nm^3/H	1.6	1	30.6	1.3	1	27.6	1.2	1	26.3
Fire Fighting System	–									
Control System	–									
Communications System	–									
Lighting & Earth	–			54.7			55.1			53.1
Flare & Blow down	–									
Common stack	–									
Auxiliary	–									
Sub-total				187.1			177.8			173.8
Total Construction Cost				922.2			950.0			1,009.8

Table 9.3 Paid-up Royalties

	Capacity per unit in	Fluid Coker Case		Eureka Case		M-DS Case	
		Capacity	Rayalties 10^6 US$	Capacity	Royalties 10^6 US$	Capacity	Royalties 10^6 US$
Fluid Coker	BPSD	43,600 X 2	*	–	–	–	–
Eureka	BPSD	–	–	42,400 X 2	*	–	–
M-DS	BPSD	–	–	–	–	40,700 X 2	*
Hydrodesulfurization	BPSD	60,900 X 2	*	16,100 X 2	*	21,400 X 2	*
				45,900 X 2	*	48,800 X 2	*
Hydrogen Generator	$MMNm^3/SD$ as H_2	0.90 X 2	0.44	1.08 X 2	0.53	1.79 X 2	0.84
Sulfur Recovery	T/SD as S	255 X 2	0.54	285 X 2	0.58	279 X 2	0.57
Flue Gas Desulfurization	10^6 Nm^3/H	1.6	1.11	1.3	1.02	1.2	0.98
Total			2.09		2.13		2.39

* Included in construction cost

Table 9.4 Initial Catalyst and Chemical Costs

	Capacity per unit in	Fluid Coker Case		Eureka Case		M-DS Case	
		Capacity	Cost 10^6 US$	Capacity	Cost 10^6 US$	Capacity	Cost 10^6 US$
Fluid Coker	BPSD	43,600 X 2	0.34	–	–	–	–
Eureka	BPSD	–	–	42,400 X 2	0.03	–	–
M-DS	BPSD	–	–	–	–	40,700 X 2	0.87
Hydrodesulfurization	BPSD	60,900 X 2	4.00	16,100 X 2 45,900 X 2	3.43	21,400 X 1 48,800 X 2	21.98
Hydrogen Generator	MMNm³/SD as H_2	0.90 X 2	1.12	1.08 X 2	1.35	1.79 X 2	2.13
Acid Gas Treating	T/SD as H_2S	67.4 X 2 22.8 X 2 191.7 X 2	0.09	315 X 2	0.10	308.8 X 2	0.10
Sulfur Recovery	T/SD as S	255 X 2	0.15	285 X 2	0.17	279 X 2	0.16
Flue Gas Desulfurization	10^6 Nm³/H	1.6	0.03	1.3	0.02	1.2	0.02
Total			5.73		5.10		25.26

Table 9.5 Pre-operating Expenses

		Fluid Coker Case	Eureka Case	M-DS Cases
		(10^6 US$)	(10^6 US$)	(10^6 US$)
1.	Operator Training	7.25	7.25	7.25
2.	Administrative Costs	3.19	3.19	3.19
3.	Startup Costs	5.96	6.20	6.20
	Total	16.40	16.64	16.64

9.2 WORKING CAPITAL

Working capital is a fund necessary for the refinery production activities. The working capital includes the following items:

9.2.1 Land Rent

If land is purchased, the cost required can be regarded as an item of fixed capital, which is not subject to depreciation, unlike the fixed-capital items described in the preceding Section. If land is on lease, the rent required can be included in the operating cost, as described later. However, land cost is not evaluated in this study and rather regarded as free of expense, as instructed by the Venezuelan side.

9.2.2 Raw Material and Product Inventories

Raw materials and products are assumed to have average inventories as much as 50% of tank storage capacities. Inventory investments have been calculated by multiplying these inventories with respective unit costs.

Unit costs are assumed to be as follows:

Raw crude oil	:	US$10/BBL
Light gas oil in mixed crude	:	Similar to the cost of raw crude oil.
Improved crude oil	:	Operating cost of the refinery (a value obtained after sales of power and fuel gas have been deducted.)

Light gas oil in the mixed crude oil is circulated between refinery and oil field, and it is regarded as non-priced. Since, however, an initial quantity of gas oil has to be purchased from outside for crude oil production, this quantity is taken in the cost calculation.

9.2.3 Catalyst and Chemical Inventories

No process for this refinery is required to replace catalysts in short periods. Thus, no inventory of catalyst is included in the working capital. Chemicals are held in stock in amounts sufficient to operate the refinery for two months. Costs of these inventories are included in the working capital.

9.2.4 Spare Parts and Warehouse Inventory

Costs of spare parts and warehouse inventories are estimated at 2% of the total construction cost. These costs are included in the working capital.

9.2.5 Cash in Hand

It is planned that two months of expenditures from direct operating cost, i.e., the operating cost excluding raw material cost and depreciation cost, is always held in cash.

9.2.6 Balance Between Accounts Receivable and Accounts Payable

A month is set as a period of grace until the refinery receives payments for its products. Thus, a month of turnover is appropriated for the account receivable. Similarly, a month is set as a period of grace for the payments of raw material costs. Thus, a month of purchase is appropriated for the account payable. The balance between the afore-mentioned account receivable and the account payable is appropriated in the working capital. Table 9.6 gives the estimated working capital amounts.

Table 9.6 Working Capitals

		Fluid Coker Case	Eureka Case	M-DS Case
		(10^6 US$)	(10^6 US$)	(10^6 US$)
1.	Land	–	–	–
2.	Oil Inventories	36.93	37.77	36.41
3.	Catalyst and Chemical Inventories	0.01	0.07	0.10
4.	Spare parts and Warehouse Supplies	18.25	19.00	20.20
5.	Cash in hand	11.23	11.11	14.99
6.	Balance of Accounts Receivable and Accounts payables	60.56	55.68	62.39
	Total	126.98	123.63	134.09

CHAPTER 10 OPERATING COSTS AND ECONOMIC ANALYSIS

This chapter deals with the calculation of operating costs required for the upgrading refinery, using the capital requirement calculated in the preceding chapter as the base, and evaluates the economics of the refinery.

10.1 OPERATING COSTS

The operating costs are grouped into direct production costs and fixed charges. Table 10.1 give a summary of operating costs. The direct operating costs are given as the operating costs from which raw material cost and depreciation cost are deducted.

In the sections below there will be described the bases and methods used to estimate respective cost items.

10.1.1 Direct Production Costs

(1) Raw material cost

The raw material cost includes the following items, which are calculated by multiplying respective unit costs with the required quantities.

Raw materials	Unit costs
Raw crude oil (in mixed crude oil)	US$10/BBL
Diluent (in mixed crude oil)	Non-priced
Natural gas (raw material for H_2)	US$3/MMBTU
Limestone	Non-priced

(2) Catalysts and chemical

With the operation going on at the refinery, it becomes necessary to make-up the catalysts and chemicals. Costs of these materials are estimated on a yearly basis, and are appropriated in the direct production cost.

(3) Utility cost

If utilities are purchased from outside, a cost must be accounted for this purpose. Such a cost may vary with the changes in the operating rate. In the case of this project, however, no utility cost is accounted. Because the refinery is planned on a utility self-sufficiency policy and industrial water, the only item to be supplied from an outside source, is regarded as free of charge.

(4) Operating supplies

Operating supplies include such items as lubricant, grease, instrumentation recording sheets, office supplies, automobile gasoline, etc. The cost for all these items is set at 0.15% of the refinery construciton cost.

Although the direct production cost may also include product packaging cost and yearly patent charges, these items are not considered under this project.

10.1.2 Fixed Charges

(1) Depreciation

Depreciation cost is calculated by applying the following depreciation method to the fixed capital obtained as in the preceding chapter.

Depreciation method:	Straight line method
Depreciation period:	16.6 years after plant start-up
Salvage value ratio:	0%

(2) Property Taxes

The property taxes imposed on fixed assets are usually appropriated in the fixed charges, but they are not in the case of this project.

(3) Insurances

These are the cost for various insurances paid after the plant goes into operation, but they are not included in the operating cost calculation of this study.

(4) Maintenance and repair cost

This item is required to proceed with smooth operation of the plant, and includes the costs for periodical maintenance, daily and periodical inspections, evaluations, accident forecasting, and small repairs. This cost is calculated totally by multiplying the construction cost with the following factors:

Process units:	4%/annum
Utility facilities:	3%/annum
Offsite facilities:	1.5%/annum

(5) Operating labor cost

This cost is required for those operators who run the refinery facilities. The number of personnel fixed as in Chapter 8 is multiplied with a unit cost of US$22/man-hour which corresponds to an average salary payment, to give this cost. The cost includes allowances other than salaries, and welfare cost, etc., in addition to salaries.

(6) Plant overhead

This cost is required for the administrative and clerical services to the refinery, and includes administrative personnel cost and office expences. The cost is totally calculated as 44% of the operating labor cost.

Apart from the foregoing direct production costs and fixed charges, a refinery necessitates general administration cost and marketing cost, if the refinery is regarded as an independent company.

Under this project, however, these costs are excluded from the operating costs as they are appropriated in the head office expenses.

Table 10.1 Operating Costs

	Fluid Coker Case	Eureka Case	M-DS Case
1. Direct Production Costs (10^6 US\$/Year)			
(1) Raw Material Cost	528.60	541.19	527.11
(2) Catalyst & Chemicals	4.41	2.37	22.68
(3) Utility Cost	–	–	–
(4) Operating Supplies	1.37	1.42	1.51
Sub Total	534.38	544.98	551.30
2. Fixed Charges (10^6 US\$/Year)			
(1) Depreciation	57.02	58.66	63.50
(2) Taxes	–	–	–
(3) Insurances	–	–	–
(4) Maintenance & Repair	29.08	30.66	33.24
(5) Operating Labor Cost	22.57	22.57	22.57
(6) Plant Overhead	9.93	9.93	9.93
Sub Total	118.60	121.52	129.24
3. Total Operating Costs (10^6 US\$/Year)	652.98	666.50	680.54
4. Direct Operating Costs (10^6 US\$/Year)	67.36	66.65	89.93
5. Direct Operating Costs Per unit crude oil			
(US\$/BBL of Mixed crude oil)	0.99	0.98	1.39
(US\$/BBL of Raw crude oil)	1.29	1.27	1.80
(US\$/BBL of Improved crude oil)	1.63	1.62	2.15

Total operating costs mean the costs from 2nd operating year until 16th operating year.

10.2 ECONOMIC ANALYSIS

The refinery economics is analyzed based on those data calculated as in Chapter 9 and Section 10.1 of this chapter. Results of this analysis will be referred as a guide to a further detailed analysis which will be carried out in the future.

The economic analysis is carried out by adding sensitivity analysis to the evaluation of base cases. The sensitivity analysis includes the following items:

Construction cost, and

Raw crude oil cost.

10.2.1 Bases and Procedures

The economic analysis is carried out on a fixed 1980 basis. The refinery is studied on the premises that it will go into operation in early 1988. No consideration is given to any escalation of costs for the period from 1980 to 1988. The analysis utilizes the calculation of return on equity (ROE) on a discounted cash flow (DCF) basis.

(1) Definition of terms

The economic analysis in this study employs the method of ROE(DCF), as designated by the Venezuelan side. The ROE(DCF) equation is given by:

$$\sum_{i=0}^{n} \frac{Ei}{(1+r)^i} = \sum_{i=0}^{n} \frac{Ai}{(1+r)^i}$$

Where r = ROE(DCF)

E_i = Equity capital investment in the i-th year.

A_i = After-tax profit plus depreciation.

n $\doteq$ Project life plus construction period expressed in years.

(2) Study bases

Premises for base cases are set as follows:

(a) Base-case refineries:

Base cases are the following three cases of plans for an Orinoco heavy oil upgrading refinery:

Fluid coker case

Eureka case

M-DS case

(b) Operating rate at the initial stage of operation:

The refinery cannot be put into full operation for a year after it is started up. It is assumed that during that year the refinery operates at a rate of 50% and goes into full operation from the second year. Full-scale operations of units and facilities are set as follows:

Process Units:	330 days of operation per year at their design capacities.
Utility facilities:	365 days of operation throughout the year at their design capacities.
Offsite facilities:	365 days of operation throught the year at their design capacities.

(c) Raw material cost

The cost of raw crude oil is set at US$10/BBL on the 1980 basis. Light gas oil used as the diluent to be mixed with raw crude oil passes through several processing steps outside the refinery, but the diluent throughput of the refinery never change; that is, the quantity of incoming diluent in the mixed crude oil is always equal to the quantity of the diluent leaving the refinery for circulation. Therefore, the diluent cost is not calculated as it is formally considered as nil. The cost of natural gas used as the raw material for hydrogen is set at US$3/MMBTU. The cost of limestone, the auxiliary raw material for gypsum production, is regarded as nil.

(d) Construction cost, operating cost, etc.

These costs are based on the calculations given in Chapter 9 and Section 10.1.

(e) Product price

The price given in Table 10.2 is used for the main product of improved crude oil (synthetic crude oil). That part of light gas oil used as the diluent is not calculated for its product cost, just as described for the raw material cost in paragraph (c), above.

By-produced sulfur and gypsum have no price. Among the utilities supplied to the outside of refinery, the products of power and fuel gas would have the following prices on the 1980 basis:

Power:	US$0.023/KWH
Fuel gas:	US$3/MMBTU

(f) Financing aspects

Capital requirement of this project is totally met by owned capital, with no loan being used.

(g) Capital expending schedule

Capital investment is all owned capital. Table 10.3 gives a capital expending schedule.

(h) Project schedule

The refinery project proceeds on the following schedule:

Completion of construction work: Late 1987

Table 10.2 Product Price

Case	Fluid Coker	Eureka	M-DS
Improved Crude Oil			
Quantity, BPSD	125,000	125,000	125,000
°API	25.7	25.0	26.1
Sulfur of Crude, wt.%	0.70	0.41	0.05
Sulfur of 650°F+, wt.%	0.73	0.60	0.024
Tia Juana Medium Crude Oil			
°API	26.4	26.4	26.4
Sulfur of Crude, wt.%	1.5	1.5	1.5
Sulfur of 650°F+, wt.%	2.3	2.3	2.3
Price, US$/BBL @ 1978	23.86	23.86	23.86
Price of Improved Crude Oil			
Gravity difference, °API	0.7	1.4	0.3
US$/BBL	-0.056	-0.112	-0.024
Sulfur difference of 650°F+, wt.%	1.57	1.7	2.276
US$/BBL	+1.125	+1.32	+2.66
Price, US$/BBL @ 1978	24.929	25.068	26.496
Price, US$/BBL @ 1980	28.54	28.70	30.34

Table 10.3 Capital Expending Schedule

	1984	1985	1986	1987	1988
1. Fixed Capital					
– Construction Costs	–	20%	50%	30%	–
– Paid-up Royalties	–	–	–	50%	50%
– Initial Catalyst and Chemical	–	–	–	100%	–
– Pre-operating Expenses	–	–	–	50%	50%
2. Working Capital					
– Oil Inventories	–	–	–	50%	50%
– Catalyst and Chemical Inventories	–	–	–	100%	–
– Spare parts and Warehouse Supplies	–	–	–	50%	50%
– Cash on Hand	–	–	–	–	100%
– Balance of Account Receivable & Account Payables	–	–	–	–	100%

Refinery start-up: Early 1988

Full-scale operation: Early 1989

(i) Project life

Project life is set at 20 years after the refinery goes into operation.

(j) Depreciation

Depreciation is as set in the preceding section.

(k) Income tax

Setting is as follows:

Tax rates: 50% and 67% (two cases)

Taxing system: Uniform method.

Tax holiday: None

(l) Dividend

Payment of any dividend is not considered in the ROE(DCF) calculations.

(m) ROE(DCF) calcualtion

ROE is calculated on the assumption that the construction will go into start in early 1984. Fiscal year closes at the end of the year for all incomes and expenditures.

(n) Sensitivity analysis

The following sensitivity analysis are carried out for the above base cases (2 income tax cases for each of the 3 process cases; hence, 6 cases in all):

- Construction cost: 20% decreased case
  20% increased case
- Raw crude oil cost: 50% decreased case (US$5/BBL)
  50% increased case (US$15/BBL)

10.2.2 Result of Study

The calculation results of ROE based on the above bases and procedures are summarized in this sub-section.

(1) Base case

The calculation results of base cases are shown on Table 10.4.

Table 10.4 Economic Analysis Summary

Case	Fluid Coker	Eureka	M-DS
	ROE, %		
Income Tax : 50% Construction cost : Base Raw Crude Oil : Base	25.0	22.9	23.1
Income Tax : 67% Construction cost : Base Raw Crude Oil : Base	18.7	17.1	17.2

(2) Sensitivity analysis

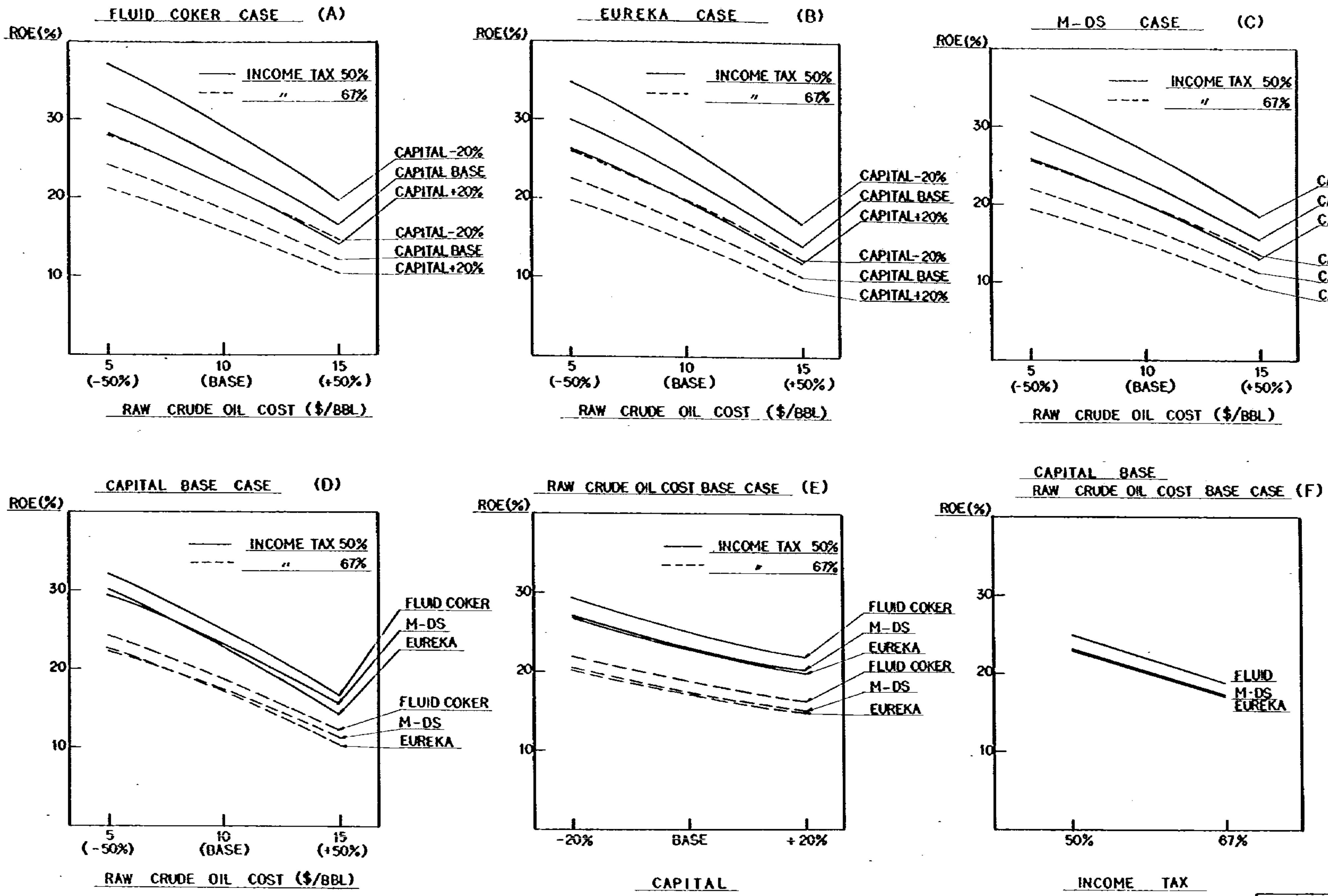
The calculation results of sensitivity analysis are shown in Table 10.5.

(3) Consideration

(a) Fluid coker case

In the fluid coker case, the relation between ROE and raw crude oil cost is shown in Fig. 10.1 (A) using parameters of construction cost and income tax.

Income tax 50% case:



Note : CAPITAL MEANS CONSTRUCTION COST.

ORINOCO HEAVY OIL UPGRADING PROJECT

RESULT OF ECONOMIC STUDY

FIG. 10.1

Table 10.5 Sensitivity Analysis Summary

Sensitivity Items: Income Tax	Construction	Raw Crude (Case)	Fluid Coker	Eureka	M-DS
			ROE: %		
50%	Base	Base	25.0	22.9	23.1
50%	Base	-50%	32.0	30.0	29.4
50%	Base	+50%	16.6	14.2	15.6
50%	-20%	Base	29.2	26.8	27.0
50%	-20%	-50%	37.1	34.9	34.2
50%	-20%	+50%	19.8	17.0	18.6
50%	+20%	Base	21.9	19.9	20.2
50%	+20%	-50%	28.3	26.4	25.9
50%	+20%	+50%	14.3	12.0	13.3
67%	Base	Base	18.7	17.1	17.2
67%	Base	-50%	24.2	22.6	22.2
67%	Base	+50%	12.2	10.3	11.4
67%	-20%	Base	21.9	20.0	20.2
67%	-20%	-50%	28.0	26.3	25.7
67%	-20%	+50%	14.6	12.5	13.7
67%	+20%	Base	16.3	14.8	15.0
67%	+20%	-50%	21.3	19.9	19.5
67%	+20%	+50%	10.4	8.7	9.7

Income Tax : Tax rates 50% and 67%

Construction : Construction cost

Base ..... study base
-20% .... 20% decreased case
+20% ... 20% increased case

Raw Crude : Raw crude oil cost

Base ..... study base (US$10/BBL)
-50% .... 50% decreased case (US$5/BBL)
+50% ..... 50% increased case (US$15/BBL)

Raw crude oil cost	ROE
50% increase	8-9% decrease
50% decrease	6-8% increase

Consturction cost	ROE
20% increase	2-4% decrease
20% decrease	3-5% increase

Income tax 67% case:

Raw crude oil cost	ROE
50% increase	6-7% decrease
50% decrease	5-6% increase

Construction cost	ROE
20 % increase	2-3% decrease
20% decrease	2-4% increase

Income tax increased case from 50% to 67%:

ROE decreases 4-9%

(b) Eureka case

In the Eureka case, the relation between ROE and raw crude oil cost is shown in Fig. 10.1 (B) using parameters of construction cost and income tax.

Income tax 50% case:

Raw crude oil cost	ROE
50% increase	8-10% decrease
50% decrease	7-8% increase

Construction cost	ROE
20% increase	2-4% decrease
20% decrease	3-5% increase

Income tax 67% case:

Raw crude oil cost	ROE
50% increase	6-8% decrease
50% decrease	5-6% increase

Construction cost	ROE
20% increase	2-3% decrease
20% decrease	2-4% increase

Income tax increased case from 50% to 67%:

ROE decreases 3-9%

(c) M-DS case

In the M-DS case, the relation between ROE and raw crude oil cost is shown in Fig. 10.1 (C) using parameters of construction cost and income tax.

Income tax 50% case:

Raw crude oil cost	ROE
50% increase	7-8% decrease
50% decrease	6-7% increase

Construction cost	ROE
20% increase	2-4% decrease
20% decrease	3-5% increase

Income tax 67% case:

Raw crude oil cost	ROE
50% increase	5-7% decrease
50% decrease	5-6% increase

Construction cost	ROE
20% increase	2-3% decrease
20% decrease	2-4% increase

Income tax increased case from 50% to 67%:

ROE decreases 4-8%

(d) ROE comparison on base case of construction cost

The relation between ROE and raw crude oil cost on the base case of construction cost is shown in Fig. 10.1 (D).

Income tax 50% case:

ROE of fluid coker case is about 2% higher than ROE of Eureka and M-DS cases.

ROE of Eureka case and M-DS case are reversed at a point of raw crude oil cost US$8/BBL.

Eureka case requires much charge of raw crude oil and its sensitivity effects to ROE.

Income tax 67% case:

Same as 50% case

(e) ROE comparison on base case of raw crude oil cost

The relation between ROE and construction cost on the base case of raw crude oil cost is shown in Fig. 10.1 (E).

Income tax 50% case:

ROE of fluid coker case is about 2% higher than ROE of Eureka and M-DS cases.

Income tax 67% case:

Same as 50% case

(f) ROE comparison on income tax sensitivity

The relation between ROE and income tax on the base cases of construciton cost and raw crude oil cost is shown in Fig. 10.1 (F).

Income tax increased case from 50% to 67%:

ROE decreases about 6%

(4) Computer output

The calculation sheets of economic analysis for the following three cases are attached in ANNEX.

	Fluid coker	Eureka	M-DS
Construction cost	base	base	base
Raw crude oil cost	base	base	base
Income tax	50%	50%	50%
Case Number on the sheet	F501	E501	M501

The calculation sheets include:

- Profit and loss (Income statement)
- Funds outlook
- Cash flow analysis

CHAPTER 11 DISCUSSION ON STUDY RESULTS

The study results were described in Chapters 5-10 on the three process cases. In this chapter there will be discussed major items of process evaluation. Discussion in this chapter excludes those items which are common to all the three cases and thus require no further study or those well-known units, such as distillation units, acid gas treatment units, sulfur recovery units, hydrogen generation plant, and general offsite facilities.

11.1 REFINERY PLAN

11.1.1 Process Scheme (Refer to Figs. 5.4, 5.5 and 5.6)

Major differences in process combinations of the three cases lie in the upgrading and hydrotreating processes. The fluid coker case and the Eureka case use the steps of:

Thermal cracking → Hydrotreating

to upgrade raw crude oil into improved crude oil. On the other hand, the M-DS case makes use of:

Deasphalting → Hydrodesulfurization/cracking

for the same purpose. In the former two cases, thermal cracking produces light cracked oil, which is treated by a hydrotreating unit capable of relatively easy operation to obtain improved crude oil, wherease in the latter case, a solvent deasphalting unit gives deasphalted oil of heavy quality (having a specific gravity of 1.0108), which is mixed with vacuum gas oil, and the mixture (having a specific gravity of 0.9944) is used as the feed to hydrocracking to obtain improved crude oil.

Hydrocracking of this mixed feed of heavy quality is operated under severe conditions than is hydrotreating in the former two cases. Whether the planned upgrading refinery is a success or a failure almost depends on smooth operation of this hydrocracking step in the latter case.

11.1.2 Yield of Improved Crude Oil (Refer to Table 5.5)

Table 11.1 summarizes the overall balances of materials and of energy for the three cases. The yield of improved crude oil and the energy yield are as follows:

	Fluid coker	Eureka	M-DS
Yield of improved crude oil			
Vol.% on raw crude oil	79.0	78.8	82.8
wt.% on raw crude oil	70.4	70.4	73.5
Calorie % on raw crude oil plus natural gas	71.4	70.7	72.8

Yield of salable products (improved crude oil, off-gas and electricity)

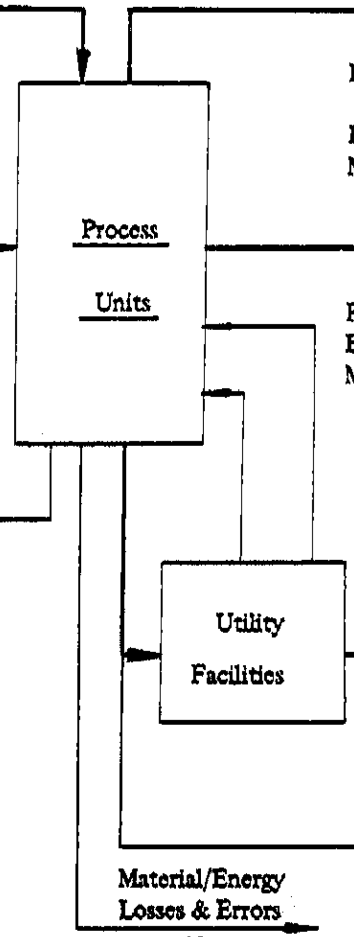
Table 11.1 Overall Material and Energy Balances

Natural Gas

	MMNm³/D	T/D	10^3 T/Y	10^{12} kcal/Y
F	0.198	156	52	0.56
E	0.518	409	135	1.46
M	0.850	671	221	2.39

Raw Crude Oil

	BPSD	T/D	10^3 T/Y	10^{12} kcal/Y
F	158,160	25,420	8,389	82.07
E	158,710	25,510	8,418	82.36
M	151,055	24,280	8,012	78.39



Off Gas to Oil Field

	MMNm³/D x D/Y	MMNm³/Y	10^{12} kcal/Y
F	1.08 x 295 0.485 x 70	352.6	4.41 –
E	–		–
M	–		

Improved Crude Oil

	BPSD	T/D	10^3 T/Y	10^{12} kcal/Y
F	125,000	17,890	5,904	59.04
E	125,000	17,970	5,930	59.30
M	125,000	17,840	5,887	58.87

Refinery & Boiler Fuel 10^{12} kcal/Y

F	10.34
E	13.26
M	12.27

Elec. to Oil Field

	KW	10^{12} kcal/Y
F	126,200	0.95
E	126,600	0.95
M	120,500	0.91

Surplus By-product

	10^3 T/Y	10^{12} kcal/Y
F	746	6.03
E	900	8.10
M	792	7.43

Material/Energy Losses & Errors

	10^{12} kcal/Y
F	1.86
E	2.21
M	1.30

Note:

(1) F — Fluid Coker Case
E — Eureka Case
M — M-DS Case

(2) Operating days/year
Refinery — 330
Boiler for Elec. Supply to Field — 365

(3) 1KW is converted to 860 kcal

(4) 10,000 kcal/kg is used as heating values of improved crude oils. (They are not indicated by the 3 groups.)

Calorie % on raw crude plus natural gas	78.0	71.9	74.0

The M-DS case gives the highest liquid yield of improved oil. However, as regards the energy yield of a sum of salable products including improved crude oil, off-gas, and electric power, the Fluid coker case gives the highest value.

11.1.3 Properties of Improved Crude Oil (Refer to Table 5.6)

	Fluid coker	Eureka	M-DS
Improved crude oil			
Gravity, °API	25.7	25.0	26.1
Sulfur, wt.%	0.7	0.41	0.05
Yield of fractions			
C_5 – 375°F, vol.%	15	Δ7.3	Δ9.5
375 – 650, "	30	32.4	34.0
650 – 1000, "	Δ55	Δ60.3	33.5
1000+, "	–	–	23.0
Sulfur cont. of fraction			
C_5 – 375°F, vol.%	Δ0.24	Δ0.09	0.01
375 – 650, "	Δ0.67	0.1	0.08
650 – 1000, "	0.73	0.6	0.02
1000+, "	–	–	0.03

Improved crude oils from these three cases meet the requirements for the specific gravity and the sulfur content of 25°API or higher and 1.0 wt.% or lower, respectively. But when yields of fractions and their sulfur content are closely scanned, some fractions are found to fail to reach the target levels (as indicated by the delta marks).

11.2 PROCESSES AND OPERATIONS

11.2.1 Fluid Coker Process

Feed oil is directly fed to the reactor to proceed with continuous cracking reaction on the fluidized bed which is formed by heated coke particles. This cracking process is already a technically established process. In addition to continuous operation, the fluid coker process has the following excellent features:

- The reactor is of an internal heating type in which no furnace is used; hence, the reactor does not require the maintenance of a furnace, such as decoking.
- The fluidized bed is heated by burning those coke particles formed by the reactor; there is no need of consuming variable liquid fuel oil.

Coker operation is easy and stable. Maintenance of units is roughly similar to ordinary cases. The reactor, the burner and the coke transport lines are lined with refractory materials. The materials get little damage, thus requiring only a little repair work in every 2 or 3 years.

An item of precaution to be taken in operation and maintenance is the clogging of units by fine coke. Toa Oil has already solved this problem.

Today, many commercial fluid cokers are in successful operation, and they are highly reliable. No problem will be posed by the installation of two 45,000 BPSD units planned.

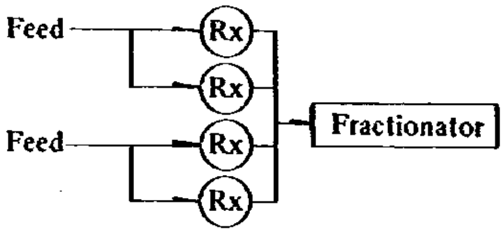
11.2.2 Eureka Process

The Eureka process is almost the same as the delayed coker process in its unit combination. The cracking unit of this process is operated in a semi-batchwise manner. Its operation is thus similar, but these processes are different in that the delayed coker uses a cracking reaction at a high temperature for a long period (on a 24-hour cycle) to carbonize the feed up to the coke. On the other hand, the Eureka process produces highly aromatic pitch by blowing a large amount of steam superheated to a high temperature into the unit to proceed with cracking and polymerizing reactions with the residual pitch, while suppressing its gasification ratio as low as possible. Precautions should be taken in the following points:

- Plugging of pipes caused by solidification of pitch.
- Fatigue in the materials of both the reactor and the flaker, due to the changes in temperature.

Troubleshooting of these difficulties are already established, and there is no special problem in operation. A commercial unit having a capacity about 20,000 BPSD is successfully in operation at Chiba Works of Eureka Industry. Another similar unit having a capacity of 20,000 BPSD is now under construction in the People's Republic of China.

It is planned in the study to adopt the following two trains of units to treat 84,656 BPSD of feed.



x 2 Trains Rx: 2 Reactors

There are two separate feed lines for 1 train, therefore, the feed quantity per feed line amounts to 21,164 BPSD. The capacity of unit is roughly on the same level as the scale of the unit now in operation. Thus, it is unlikely that such a capacity may cause any problem.

11.2.3 M-DS Process

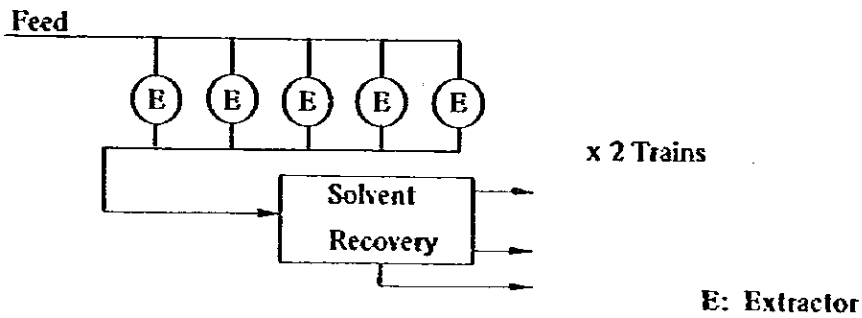
Maruzen Oil has designed an extractor having a special internal mechanism, and developed a process for effectively separating oil fractions, asphaltene and metals from heavy residues at a low solvent ratio. A 24-BPSD pilot plant based on this process is now in operation for a research purpose. Care should be taken in operation on the following points:

- Pipe clogging due to asphalt solidification inside the pipes.
- Contamination of solvent lines caused by entrained heavy oils.

In spite of the requirement for precautions on the above points, similar types of many solvent deasphalting units are in operation on commercial scales, and they pose no special problem of operation. Examples of other solvent deasphalting units in commercial operation are as follows:

- A 13,500 BPSD SDA/demetalizing unit at Corpus Christi Refinery of Champlin Oil Co.
- A 60,000 BPSD unit at Richmond Refinery of Chevron Oil of the U.S.

It is planned in the study to adopt the following two trains of units to treat 81,283 BPSD of feed.



Extractor 4,100ϕ X 23,000 mm, 8,128 BPSD/Extractor

Five extractors are arranged in parallel for a train.

11.2.4 Yields of Liquid Oil Fractions from the Three Upgrading Processes

Refer to Table 5.2, 5.3, and 5.4.

	Yield of liquid oil on feed
Fluid coker cracked oil:	62.48 vol.% (53.35 wt.%)
Eureka cracked oil:	68.85 vol.% (59.00 wt.%)
M-DS DAO:	68.57 vol.% (65.33 wt.%)

The M-DS process gives the highest yield, followed by Eureka and Fluid coker in the decreasing order. It should be noted, however, that M-DS DAO is of heavy quality, requiring to use hydrocracking in the subsequent step. Liquid yield of the Fluid coker is lower than that of the Eureka process because of high gas yield of the Fluid coker. This gas is used as a supplementary boiler fuel. Any surplus amount will be supplied to the oil field.

11.2.5 Hydrotreating Units

The GO hydrodesulfurization unit used in the M-DS case, is aimed at desulfurizing a feed of only light straightrun oil. Its operation is easy and stable.

The hydrotreating unit in both of the Fluid coker case and the Eureka case is used to saturate the mixture of low-metal cracked oil and straight-run gas oils, while at the same time desulfurizing it. Its operation, too, is easy and stable. These processes are all based on established technologies. The catalysts used remain active for 1 to 3 years and pose no problem.

11.2.6 Heavy Oil Hydrodesulfurization Unit Used in the M-DS Case (VGO/DAO HDS in the M-DS case)

The feed oil of VGO/DAO mixture to be treated by the hydrodesulfurization unit has a specific gravity as quite high as 10.8° API. In order to achieve a target Sp. Gr. level of 25°API set for improved crude oil, attention will be paid to the design and operation conditions of hydrocracking. From its experiences in the hydrodesulfurization of the atmospheric residue of Middle East crude oil, Maruzen Oil has set the following conditions.

(1) Operating conditions

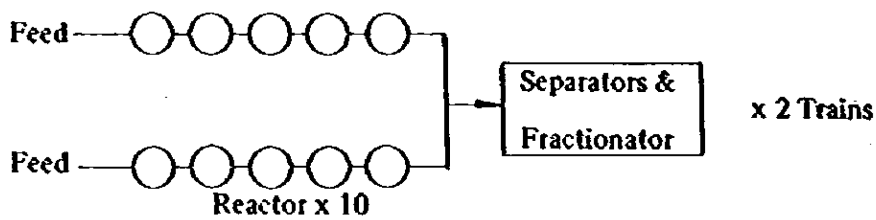
Space velocity (LHSV):	0.15 v/h/v
H_2 consumption, chem.:	1,030 SCF/B

It is noteworthy that the space velocity is significantly lower than in ordinary hydrodesulfurization. Form this space velocity we understand Maruzen's intention to allow the cracking reaction to proceed to a larger extent by taking a long contact time.

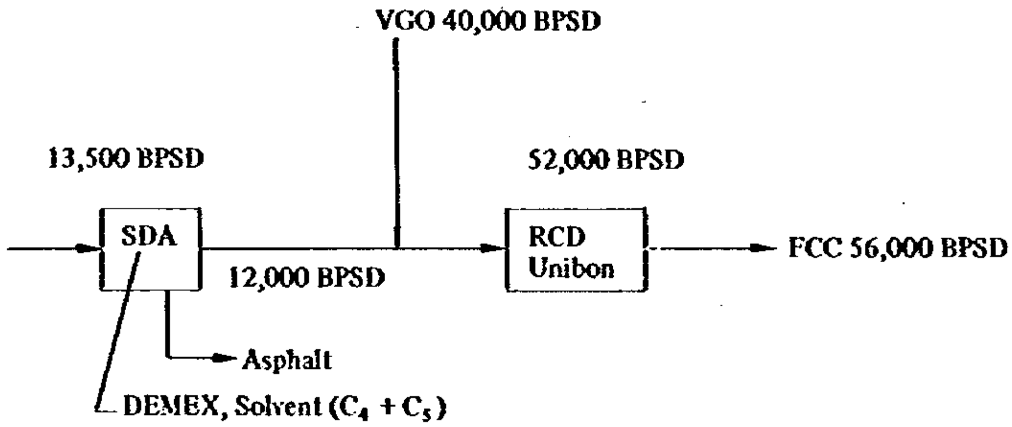
(2) Schematic flow of units

It is planned in the study to adopt the following two trains of units to treat 98,537 BPSD of feed.

Because of a low space velocity there are used as many as 10 reactors for 1 train. There are two separate parallel lines for 1 train, each of which has five reactors, as schematically illustrated below.



Based on the above conditions and unit arrangement, Maruzen gave the estimated yield and properties of the improved crude oil. These estimated results seem to allow some room for further study on the following points:
For one thing, severe cracking is required to upgrade the feed by as large as 14.2°API from 10.8 to 25.0 in terms of specific gravity. It becomes crucial, therefore, to select an optimum hydrodesulfurization/cracking unit capable of affording this extent of upgrading, while suppressing gas evolution to a low level, as set by Maruzen Oil.
With regard to the content of metals, the feed oil (VGO/DAO mixture) is similar to the atmospheric residue of Kuwait crude or Khafji crude. In this regard, the afore-mentioned operating conditions are unlikely to give significant damages to catalyst activity.
As an example of VGO/DAO processing, there may be mentioned Corpus Christi Refinery of Champlin Oil Co.



As seen in the sketch above, 12,000 BPSD of DAO is mixed with as much as 40,000 BPSD of VGO, and the mixture is used as the feed to RCD Unibon. No information is available on process performance of this refinery.
It is tentatively assumed that the construction periods described in Chapter 8 and 10 are same for all of the three cases.
Actually, however, the construciton period in the M-DS case may be relatively longer than in other cases, as far as major refinery equipment is concerned. For example, the VGO/DAO hydrodesulfurization units used in M-DS have 20 reactors of a 500-ton class having large wall thickness. If high-pressure separators are taken into account, the number of large size units is exceptionally large for a single project. The manufacturers of these types of equipment are limited in their number and supply capacity. They are, for example, Creusot-Loire (France), Nuovo Pignone S.P.A. (Italy), Babcock & Wilcox (the U.K. and the U.S.), Nooter Corp. (the U.S.), The Japan Steel Works (Japan), Kobe Steel (Japan) and MHI (Japan).
Dependent of the timing of project, the procurement of these types of machinery

may pose problems, such as extended construction period, cost increases, etc. Care should be taken fully to practical work problems, such as land transportation of these types of equipment.

11.2.7 Hydrogen Consumption by the Hydrotreating Process (Refer to Table 5.8)

Case	Unit	H_2 Consumption
Fluid Coker	HTR	547 SCF/B Feed
Eureka	#1, HTR	440 "
	#2, HTR	680 "
M-DS	GO HDS	401 "
	VGO/DAO HDS	1,193 "

The above levels of hydrogen consumption are considered reasonable for the most part, when they are evaluated from the feed properties and the reaction conditions.

11.3 RESIDUAL PRODUCTS

All the three cases produce quite large amounts of residuals (See Table 5.5). It is premised, as base of the study, that either residual shall be utilized as the fuel for industrial boilers. The residuals obtained in the three cases are fluid coke, Eureka pitch, and M-DS asphalt. All of them are solid at normal temperature, and contain high levels of sulfur and metals. (See Table 7.1).

11.3.1 Combustion Systems

(1) Fluid coke: Pulverized fuel combustion.

Fluid coke shows a Hardgrove Index of 30 or less. It is harder than ordinary coal, but it can be pulverized easily. Coke is finely pulverized by mills to a particle size of 200 mesh or higher (80% pass through 200-mesh screen). Fine coke thus obtained is sent together with air to the burner for combustion.

Fluid coke has an ignition temperature of about 870°C. Its inflammability is remarkable at a high temperature. In order to avoid unstable combustion caused by the high ignition temperature, the burner requires a supplementary fuel in an amount of about 10% of the total heat quantity. Fine coke is usually atomized downward in a U-type firing manner.Fluid coker boilers are already in commercial operation for the power generation purpose. When this case is adopted, there will be no problem to be concerned with. A thermal efficiency of about 85% can be expected.

(2) Eureka pitch: Pulverized fuel combustion.

Eureka pitch has a softening point of 220°C and a volatile matter of 45.3%. It has a lower ignition temperature and better inflammability than fluid coke. As a brittle solid, the pitch can be crushed easily by a hammer mill, according to the licensor. Good inflammability makes Eureka pitch fully usable as a boiler fuel on a commercial scale. Combustion requires no supplementary fuel.

(3) M-DS asphalt: High-temperature liquid combustion

Maruzen Oil proposes a system of burning M-DS asphalt in the liquid form at a high temperature, using conventional steam-atomizing type of burners. This system is not so easy, as compared with the ordinary cases of heavy oil combustion by means of similar boilers. In the past, there was a case where asphalt was used for industrial boilers, but in that case the asphalt used had a pour point of 50°C or less.

Asphalt handling was far easier than the handling of M-DS asphalt which has a softening point of about 162°C.

Maruzen Oil has the opinion that the asphalt of 360°C and 100 cst can be atomized and burnt by blowing steam superheated to as high as 500°C into the inside of the burner. No technical data has been reported so far on the combustion of M-DS asphalt or similar substances. Therefore, if this system is to be put to practical use, careful commercialization tests will be required to investigate this combustion system and the mechanism of its fuel piping system. Major items of study in commercialization are as follows:

(a) Mechanism of atomization
(b) Keeping all fuel lines at high temperatures.
(c) Replacement of asphalt with light oil in fuel lines.
(d) Coke deposit on the burner tip.
(e) Safety of fuel lines.

Meanwhile, JICA conducted a pulverization test using BDA asphalt (See Chap. 7, Volume II) which has similar properties to M-DS asphalt, with a view to utilizing it for solids combustion. The test turned out to be a failure, because a sticky material was formed in the ball mill, and its deposition caused the mill to fall into troubles. As a result, it has been concluded that M-DS asphalt and similar material is difficult to be pulverized fuel combustion method. If, however, a particle size of 1-7 mm is permitted, pulverization seemed to be practicable, judging from the situations in this pulverizing test.

Commercialization of a fluidized-bed combustion boiler (FBC boiler) is anticipated several years ahead from now. If it is completed, M-DS asphalt in the solid form will become a promising useful fuel.

11.3.2 Power and Steam Supplies

A boiler plant has been planned for each of the three schemes, assuming that the proposed combustion methods described in the preceding subsection 11.3.1 can be put into practical use.

(1) Requirements of power and steam

The requirements of electric power and steam for normal operation are as follows:

	Fluid coker	Eureka	M-DS
Electric power, KW	200,400	208,000	203,000
for oil field, "	126,200	126,600	120,500
for refinery, "	74,200	81,400	82,500
Steam, for refinery T/H (process use only)	323.7	416.0	485.8

(2) Number and capacity of boilers

It has been assumed that the boilers would be in operation for 365 days a year and that a spare unit is installed.

	Fluid coker	Eureka	M-DS
CO Boiler, 50 kg/cm^2G	200T/H X 2	–	–
Boiler, 100 kg/cm^2G	260T/H X 4	240T/H X 6	240 T/H X 6
Power generation	18 MW X 2	–	–
	55 MW X 4	46 MW X 6	44 MW X 6

The CO boilers use, as its fuel, a mixture of CO gas from the coker burner and refinery off-gas. These boilers are located separately from the coke boilers.

(3) Consumption of by-produced residual fuel
(See Table 7.1)

	Fluid coker	Eureka	M-DS
Residuals, Consumption 10^3 T/Y	545	839	780
Residuals, Produced 10^3 T/Y	1,291	1,739	1,572
Residuals, Surplus 10^3 T/Y	746	900	792

As obvious from the above table, all the three cases give surplus by-products in amount more than 50% of their production.

(4) Other items

– Flue gas desulfurization:– A wet type recovery method has been adopted in this study, wherein lime slurry is used to recovery sulfur from flue gas at a recovery rate of 90% (See Table 6.9).

– High-temperature corrosion (Anti-vanadium corrosion measures):– Corrosion caused by vanadium at high temperatures can be prevented by the use of magnesium hydroxide, ($Mg(OH)_2$), and/or the optimum arrangement of superheated-steam pipings.

- Low-temperature corrosion (Anti-sulfur corrosion measure):– As a counter-measure against low-temperature corrosion caused by sulfur, the feed water temperature is set at as high as 190°C.

11.3.3 Utilization of Surplus Residuals

If a power and steam generation plant is planned in accordance with the premises set for by-product use, each of the three process schemes produces a large surplus, as given in the table below.

	Fluid coker	Eureka	M-DS
Surplus Residuals T/D	2,262	2,727	2,400
Surplus Residuals T/Y	746,345	899,992	792,126

Fluid coke, Eureka pitch or M-DS asphalt is produced in quite a large surplus. This valuable fuel resource cannot be left without any planning on its utilization. This surplus residual is not studied, however this matter is one of important items to be reviewed prior to the realization of the project. Utilization of surplus by-product will have to be decided, inconsideration of the future prospect as to how the demands will develop for power, fuel gas, or gas for chemical use in the Cerro Negro area. The surplus by-product should be, in principle, consumed in the area.

If this resource can be priced low, its economically effective utilization may be found in the following fields, depending on its demand:

(1) Conversion to electric power, which is supplied to EDELCA.

(2) Conversion to fuel gas and feed gas for chemical plant; and the recovery of vanadium (V) and nickel (Ni).

(3) Use as a special binding material of steelmaking coke.

11.4 ECONOMICS

11.4.1 Capital Requirement (See Table 9.1)

The capital requirement for the upgrading refinery is as given in the table below. It should be noted that this requirement is estimated on the 1980 basis. In order to estimate budget for the planned schedule, the capital requirement will have to be escalated in response to the time.

	Fluid Coker	Eureka	M-DS
Total Capital Requirements (10^6 US$)	1,073.40	1,097.50	1,188.18
Cost reference	100	102	111

The capital requirement of the Fluid coker case is almost the same as that of the Eureka case, whereas that of the M-DS case is 10% higher. This is because the VGO/DAO HDS process is much costly, despite the fact that the M-DS process itself is relatively less expensive than Fluid coker and Eureka processes.

11.4.2 Operating Cost (See Table 10.1)

The total operating costs of the upgrading refinery are as follows:

	Fluid coker	Eureka	M-DS
Total Operating Costs (10^6 US$/Y)	652.98	666.50	680.54
Total Operating costs per unit improved crude oil (US$/BBL)	15.83	16.16	16.50
Direct Operating Costs (10^6 US$/Y)	67.36	66.65	89.93
Direct Operating Costs per unit improved crude oil (US$/BBL)	1.63	1.62	2.15

There is no large difference in the total operating costs among the three cases. As for the direct operating cost, the M-DS case is most expensive. This is because of a high construction cost and because the cost of catalyst used in the VGO/DAO HDS process pushes up the direct operating cost significantly.

11.4.3 Economic Analysis (See Table 10.4)

The following ROE Values are given by the DCF calculations in the base case of an income tax rate of 50%.

	Fluid coker	Eureka	M-DS
ROE	25.0%	22.9%	23.1%

The Fluid coker case gives the best ROE, owing to the sales of fuel gas, ROE level of the Eureka case is roughly the same as that of the M-DS case.

The M-DS case gives a good ROE level in spite of its somewhat higher capital requirement and operating cost. This is because its improved crude oil having the lowest sulfur content can be favorably priced.

ANNEX

CALCULATION SHEET OF ECONOMIC ANALYSIS

- CASE NO. F501 Fluid Coker Case
- CASE NO. E501 Eureka Case
- CASE NO. M501 M-DS Case

ANNEX

ORINOCO HVY-O PJ

ITEM NO. BASEBASE

CASE NO. F501

E C O N O M I C

E V A L U A T I O N

P R O G R A M

JOB NO.

DATE 19800721

ENGINEER T.SHIMOG

PROFIT & LOSS

10**3DOLL F501

	1 1984	2 1985	3 1986	4 1987	5 1988	6 1989	7 1990	8 1991	9 1992	10 1993
* REVENUE										
IMP. CRUDE	0	0	0	0	588637	1177273	1177273	1177273	1177273	1177273
ELECTRIC POWER	0	0	0	0	12713	25427	25427	25427	25427	25427
FUEL GAS	0	0	0	0	26300	52600	52600	52600	52600	52600
INT. RECEIVED	0	0	0	0	0	0	0	0	0	0
TOTAL	0	0	0	0	627650	1255299	1255299	1255299	1255299	1255299
* EXPENSES										
RAW MATERIAL	0	0	0	0	264298	528597	528597	528597	528597	528597
CAT.&CHEM.	0	0	0	0	4410	4410	4410	4410	4410	4410
UTILITIES	0	0	0	0	0	0	0	0	0	0
OPERATIG SUPPLY	0	0	0	0	1369	1369	1369	1369	1369	1369
TAXES	0	0	0	0	0	0	0	0	0	0
INSURANCES	0	0	0	0	0	0	0	0	0	0
MAINTE.& REPAIR	0	0	0	0	29080	29080	29080	29080	29080	29080
OPERATING LABOR	0	0	0	0	22572	22572	22572	22572	22572	22572
PLANT OH	0	0	0	0	9932	9932	9932	9932	9932	9932
DEPRECIATION	0	0	0	0	57016	57016	57016	57016	57016	57016
INT. PAID	0	0	0	0	0	0	0	0	0	0
TOTAL	0	0	0	0	388677	652975	652975	652975	652975	652975
* PROFIT BEF. TAX	0	0	0	0	238974	602324	602324	602324	602324	602324
* INCOME TAX	0	0	0	0	119487	301162	301162	301162	301162	301162
* PROFIT AFT. TAX	0	0	0	0	119487	301162	301162	301162	301162	301162
* CUM. PROFIT	0	0	0	0	119487	420649	721811	1022973	1324135	1625297

- Continued -

PROFIT & LOSS

10**03DOLL

F501

	11	12	13	14	15	16	17	18	19	20
	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003
* REVENUE										
IMP. CRUDE	1177273	1177273	1177273	1177273	1177273	1177273	1177273	1177273	1177273	1177273
ELECTRIC POWER	25427	25427	25427	25427	25427	25427	25427	25427	25427	25427
FUEL GAS	52600	52600	52600	52600	52600	52600	52600	52600	52600	52600
INT. RECEIVED	0	0	0	0	0	0	0	0	0	0
TOTAL	1255299	1255299	1255299	1255299	1255299	1255299	1255299	1255299	1255299	1255299
* EXPENSES										
RAW MATERIAL	528597	528597	528597	528597	528597	528597	528597	528597	528597	528597
CAT.&CHEM.	4410	4410	4410	4410	4410	4410	4410	4410	4410	4410
UTILITIES	0	0	0	0	0	0	0	0	0	0
OPERAT'G SUPPLY	1369	1369	1369	1369	1369	1369	1369	1369	1369	1369
TAXES	0	0	0	0	0	0	0	0	0	0
INSURANCES	0	0	0	0	0	0	0	0	0	0
MAINTE.& REPAIR	29080	29080	29080	29080	29080	29080	29080	29080	29080	29080
OPERATING LABOR	22572	22572	22572	22572	22572	22572	22572	22572	22572	22572
PLANT OH	9932	9932	9932	9932	9932	9932	9932	9932	9932	9932
DEPRECIATION	57016	57016	57016	57016	57016	57016	57016	57016	57016	57016
INT. PAID	0	0	0	0	0	0	0	0	0	0
TOTAL	652975	652975	652975	652975	652975	652975	652975	652975	652975	652975
* PROFIT BEF. TAX	602324	602324	602324	602324	602324	602324	602324	602324	602324	602324
* INCOME TAX	301162	301162	301162	301162	301162	301162	301162	301162	301162	301162
* PROFIT AFT. TAX	301162	301162	301162	301162	301162	301162	301162	301162	301162	301162
* CUM. PROFIT	1926459	2227621	2528783	2829945	3131107	3432269	3733431	4034593	4335755	4636917

- Continued -

P R O F I T & L O S S

10**03DOLL F501

	21 2004	22 2005	23 2006	24 2007	SUM
* REVENUE					
IMP. CRUDE	1177273	1177273	1177273	1177273	22956768
ELECTRIC POWER	25427	25427	25427	25427	495822
FUEL GAS	52600	52600	52600	52600	1025700
INT. RECEIVED	0	0	0	0	0
TOTAL	1255299	1255299	1255299	1255299	24478304
* EXPENSES					
RAW MATERIAL	528597	528597	528597	528597	10307635
CAT.&CHEM.	4410	4410	4410	4410	88200
UTILITIES	0	0	0	0	0
OPERAT'G SUPPLY	1369	1369	1369	1369	27376
TAXES	0	0	0	0	0
INSURANCES	0	0	0	0	0
MAINTE.& REPAIR	29080	29080	29080	29080	581600
OPERATING LABOR	22572	22572	22572	22572	451440
PLANT OH	9932	9932	9932	9932	198634
DEPRECIATION	34209	0	0	0	946459
INT. PAID	0	0	0	0	0
TOTAL	650169	595960	595960	595960	12601346
* PROFIT BEF. TAX	625130	659340	659340	659340	11876967
* INCOME TAX	312565	329670	329670	329670	5938475
* PROFIT AFT. TAX	312565	329670	329670	329670	5938489
* CUM. PROFIT	4949482	5279151	5608820	5938489	

- Continued -

FUNDS OUTLOOK

10**03DULL F501

	1	2	3	4	5	6	7	8	9	10
	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993
* CASH OUT										
INVESTMENT	0	184450	461120	319260	108610	0	0	0	0	0
CONST. COSTS	0	184450	461120	276670	0	0	0	0	0	0
PAIDUP ROYALTY	0	0	0	1050	1040	0	0	0	0	0
INIT.CAT&CHEM.	0	0	0	5730	0	0	0	0	0	0
PRODPE. EXPNS.	0	0	0	8200	8200	0	0	0	0	0
OIL INV.	0	0	0	18470	18460	0	0	0	0	0
CAT&CHEM INV.	0	0	0	10	0	0	0	0	0	0
SPARE&WAREHOUS	0	0	0	9130	9120	0	0	0	0	0
CASH ON HAND	0	0	0	0	11230	0	0	0	0	0
A/C REC.-PAY.	0	0	0	0	60560	0	0	0	0	0
REFUND OF LOANS	0	0	0	0	0	0	0	0	0	0
EQUITY	0	0	0	0	0	0	0	0	0	0
DIVIDEND	0	0	0	0	0	0	0	0	0	0
TOTAL	0	184450	461120	319260	108610	0	0	0	0	0
* CASH IN										
RETAINED EARN.	0	0	0	0	176503	358178	358178	358178	358178	358178
PROFIT AFT TAX	0	0	0	0	119487	301162	301162	301162	301162	301162
DEPRECIATION	0	0	0	0	57016	57016	57016	57016	57016	57016
LOANS	0	184450	461120	319260	0	0	0	0	0	0
EQUITY	0	184450	461120	319260	0	0	0	0	0	0
CAPITAL	0	0	0	0	0	0	0	0	0	0
TOTAL	0	184450	461120	319260	176503	358178	358178	358178	358178	358178
* SURPLUS	0	0	0	0	67893	358178	358178	358178	358178	358178
* CUM. SURPLUS	0	0	0	0	67893	426070	784248	1142425	1500602	1858779
* BALANCE OF LOANS	0	184450	645570	964830	964830	964830	964830	964830	964830	964830
EQUITY	0	184450	645570	964830	964830	964830	964830	964830	964830	964830

- Continued -

F U N D S   O U T L O O K

10**03DOLL F501

	11 1994	12 1995	13 1996	14 1997	15 1998	16 1999	17 2000	18 2001	19 2002	20 2003
* CASH OUT										
INVESTMENT	0	0	0	0	0	0	0	0	0	0
CONST. COSTS	0	0	0	0	0	0	0	0	0	0
PAIDUP ROYALTY	0	0	0	0	0	0	0	0	0	0
INIT.CAT&CHEM.	0	0	0	0	0	0	0	0	0	0
PREOPE. EXPNS.	0	0	0	0	0	0	0	0	0	0
OIL INV.	0	0	0	0	0	0	0	0	0	0
CAT&CHEM INV.	0	0	0	0	0	0	0	0	0	0
SPARE&WAREHOUS	0	0	0	0	0	0	0	0	0	0
CASH ON HAND	0	0	0	0	0	0	0	0	0	0
A/C REC.-PAY.	0	0	0	0	0	0	0	0	0	0
REFUND OF LOANS	0	0	0	0	0	0	0	0	0	0
EQUITY	0	0	0	0	0	0	0	0	0	0
DIVIDEND	0	0	0	0	0	0	0	0	0	0
TOTAL	0	0	0	0	0	0	0	0	0	0
* CASH IN										
RETAINED EARN.	358178	358178	358178	358178	358178	358178	358178	358178	358178	358178
PROFIT AFT TAX	301162	301162	301162	301162	301162	301162	301162	301162	301162	301162
DEPRECIATION	57016	57016	57016	57016	57016	57016	57016	57016	57016	57016
LOANS	0	0	0	0	0	0	0	0	0	0
EQUITY	0	0	0	0	0	0	0	0	0	0
CAPITAL	0	0	0	0	0	0	0	0	0	0
TOTAL	358178	358178	358178	358178	358178	358178	358178	358178	358178	358178
* SURPLUS	358178	358178	358178	358178	358178	358178	358178	358178	358178	358178
* CUM. SURPLUS	2216956	2575133	2933310	3291487	3649664	4007841	4366018	4724195	5082372	5440549
* BALANCE OF LOANS	964830	964830	964830	964830	964830	964830	964830	964830	964830	964830
EQUITY	964830	964830	964830	964830	964830	964830	964830	964830	964830	964830

- Continued -

F U N D S   O U T L O O K

10**03DOLL

F501

	21 2004	22 2005	23 2006	24 2007	SUM
* CASH OUT					
INVESTMENT	0	0	0	0	1073440
CONST. COSTS	0	0	0	0	922240
PAIDUP ROYALTY	0	0	0	0	2090
INIT.CAT&CHEM.	0	0	0	0	5730
PREOPE. EXPNS.	0	0	0	0	16400
OIL INV.	0	0	0	0	36930
CAT&CHEM INV.	0	0	0	0	10
SPARE&WAREHOUS	0	0	0	0	18250
CASH ON HAND	0	0	0	0	11230
A/C REC.-PAY.	0	0	0	0	60560
REFUND OF LOANS	0	0	0	0	0
EQUITY	0	0	0	0	0
DIVIDEND	0	0	0	0	0
TOTAL	0	0	0	0	1073440
* CASH IN					
RETAINED EARN.	346775	329670	329670	329670	6884940
PROFIT AFT TAX	312565	329670	329670	329670	5938489
DEPRECIATION	34209	0	0	0	946459
LOANS	0	0	0	0	964830
EQUITY	0	0	0	0	964830
CAPITAL	0	0	0	0	0
TOTAL	346775	329670	329670	329670	7849768
* SURPLUS	346775	329670	329670	329670	6776330
* CUM. SURPLUS	5787323	6116992	6446661	6776330	
* BALANCE OF LOANS	964830	964830	964830	964830	
EQUITY	964830	964830	964830	964830	

- Continued -

CASH FLOW ANALYSIS

10**03DOLL F501

YEAR	INVEST--MENT (-)	PROFIT AFTER TAX (+)	DEPREC--IATION (+)	CASH FLOW	D.C.F.	DISCOUNT RATE
1984	0	0	0	0	0	1.0000
1985	184450	0	0	-184450	-147587	0.8001
1986	461120	0	0	-461120	-295225	0.6402
1987	319260	0	0	-319260	-163551	0.5123
1988	108610	119487	57016	67893	27829	0.4099
1989	0	301162	57016	358178	117476	0.3280
1990	0	301162	57016	358178	93998	0.2624
1991	0	301162	57016	358178	75212	0.2100
1992	0	301162	57016	358178	60181	0.1680
1993	0	301162	57016	358178	48154	0.1344
1994	0	301162	57016	358178	38530	0.1076
1995	0	301162	57016	358178	30830	0.0861
1996	0	301162	57016	358178	24668	0.0689
1997	0	301162	57016	358178	19738	0.0551
1998	0	301162	57016	358178	15793	0.0441
1999	0	301162	57016	358178	12637	0.0353
2000	0	301162	57016	358178	10112	0.0282
2001	0	301162	57016	358178	8091	0.0226
2002	0	301162	57016	358178	6474	0.0181
2003	0	301162	57016	358178	5180	0.0145
2004	0	312565	34209	346775	4013	0.0116
2005	0	329670	0	329670	3052	0.0093
2006	0	329670	0	329670	2442	0.0074
2007	0	329670	0	329670	1954	0.0059
SUM	1073440	5938489	946459	5811503	0	

RATE OF RETURN 24.98 %

PAYOUT PERIOD 3.12 YEAR

ORINOCO HVY-O PJ

ITEM NO. BASEBASE

CASE NO. E501

E C O N O M I C

E V A L U A T I O N

P R O G R A M

JOB NO.

DATE 19800722

ENGINEER T.SHIMOG

P R O F I T & L O S S

10**03DOLL — E501

	1	2	3	4	5	6	7	8	9	10
	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993
* REVENUE										
IMP. CRUDE	0	0	0	0	591938	1183876	1183876	1183876	1183876	1183876
ELECTRIC POWER	0	0	0	0	12754	25507	25507	25507	25507	25507
FUEL GAS	0	0	0	0	0	0	0	0	0	0
INT. RECEIVED	0	0	0	0	0	0	0	0	0	0
TOTAL	0	0	0	0	604692	1209383	1209383	1209383	1209383	1209383
* EXPENSES										
RAW MATERIAL	0	0	0	0	270595	541189	541189	541189	541189	541189
CAT.&CHEM.	0	0	0	0	2370	2370	2370	2370	2370	2370
UTILITIES	0	0	0	0	0	0	0	0	0	0
OPERAT'G SUPPLY	0	0	0	0	1425	1425	1425	1425	1425	1425
TAXES	0	0	0	0	0	0	0	0	0	0
INSURANCES	0	0	0	0	0	0	0	0	0	0
MAINTE.& REPAIR	0	0	0	0	30360	30360	30360	30360	30360	30360
OPERATING LABOR	0	0	0	0	22572	22572	22572	22572	22572	22572
PLANT OH	0	0	0	0	9932	9932	9932	9932	9932	9932
DEPRECIATION	0	0	0	0	58663	58663	58663	58663	58663	58663
INT. PAID	0	0	0	0	0	0	0	0	0	0
TOTAL	0	0	0	0	395916	666511	666511	666511	666511	666511
* PROFIT BEF. TAX	0	0	0	0	208775	542872	542872	542872	542872	542872
* INCOME TAX	0	0	0	0	104388	271436	271436	271436	271436	271436
* PROFIT AFT. TAX	0	0	0	0	104388	271436	271436	271436	271436	271436
* CUM. PROFIT	0	0	0	0	104388	375824	647260	918696	1190131	1461567

- Continued -

P R O F I T & L O S S

10**03DOLL E501

	11	12	13	14	15	16	17	18	19	20
	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003
* REVENUE										
IMP. CRUDE	1183876	1183876	1183876	1183876	1183876	1183876	1183876	1183876	1183876	1183876
ELECTRIC POWER	25507	25507	25507	25507	25507	25507	25507	25507	25507	25507
FUEL GAS	0	0	0	0	0	0	0	0	0	0
INT. RECEIVED	0	0	0	0	0	0	0	0	0	0
TOTAL	1209383	1209383	1209383	1209383	1209383	1209383	1209383	1209383	1209383	1209383
* EXPENSES										
RAW MATERIAL	541189	541189	541189	541189	541189	541189	541189	541189	541189	541189
CAT.&CHEM.	2370	2370	2370	2370	2370	2370	2370	2370	2370	2370
UTILITIES	0	0	0	0	0	0	0	0	0	0
OPERAT'G SUPPLY	1425	1425	1425	1425	1425	1425	1425	1425	1425	1425
TAXES	0	0	0	0	0	0	0	0	0	0
INSURANCES	0	0	0	0	0	0	0	0	0	0
MAINTE.& REPAIR	30360	30360	30360	30360	30360	30360	30360	30360	30360	30360
OPERATING LABOR	22572	22572	22572	22572	22572	22572	22572	22572	22572	22572
PLANT OH	9932	9932	9932	9932	9932	9932	9932	9932	9932	9932
DEPRECIATION	58663	58663	58663	58663	58663	58663	58663	58663	58663	58663
INT. PAID	0	0	0	0	0	0	0	0	0	0
TOTAL	666511	666511	666511	666511	666511	666511	666511	666511	666511	666511
* PROFIT BEF. TAX	542872	542872	542872	542872	542872	542872	542872	542872	542872	542872
* INCOME TAX	271436	271436	271436	271436	271436	271436	271436	271436	271436	271436
* PROFIT AFT. TAX	271436	271436	271436	271436	271436	271436	271436	271436	271436	271436
* CUM. PROFIT	1733003	2004439	2275875	2547311	2818747	3090183	3361619	3633055	3904491	4175927

- Continued -

P R O F I T & L O S S

10**03DOLL

E501

	21 2004	22 2005	23 2006	24 2007	SUM
* REVENUE					
IMP. CRUDE	1183876	1183876	1183876	1183876	23085552
ELECTRIC POWER	25507	25507	25507	25507	497394
FUEL GAS	0	0	0	0	0
INT. RECEIVED	0	0	0	0	0
TOTAL	1209383	1209383	1209383	1209383	23582928
* EXPENSES					
RAW MATERIAL	541189	541189	541189	541189	10553181
CAT.&CHEM.	2370	2370	2370	2370	47400
UTILITIES	0	0	0	0	0
OPERAT'G SUPPLY	1425	1425	1425	1425	28498
TAXES	0	0	0	0	0
INSURANCES	0	0	0	0	0
MAINTE.& REPAIR	30360	30360	30360	30360	607200
OPERATING LABOR	22572	22572	22572	22572	451440
PLANT OH	9932	9932	9932	9932	198634
DEPRECIATION	35198	0	0	0	973809
INT. PAID	0	0	0	0	0
TOTAL	643046	607848	607848	607848	12860167
* PROFIT BEF. TAX	566337	601535	601535	601535	10722797
* INCOME TAX	283168	300767	300767	300767	5361396
* PROFIT AFT. TAX	283169	300768	300768	300768	5361397
* CUM. PROFIT	4459096	4759863	5060630	5361397	

- Continued -

FUNDS OUTLOOK

10**03DOLL

E501

	1 1984	2 1985	3 1986	4 1987	5 1988	6 1989	7 1990	8 1991	9 1992	10 1993
* CASH OUT										
INVESTMENT	0	189990	474970	327930	104550	0	0	0	0	0
CONST. COSTS	0	189970	474970	284980	0	0	0	0	0	0
PAIDUP ROYALTY	0	0	0	1070	1060	0	0	0	0	0
INIT.CAT&CHEM.	0	0	0	5100	0	0	0	0	0	0
PREOPE. EXPNS.	0	0	0	8320	8320	0	0	0	0	0
OIL INV.	0	0	0	18890	18880	0	0	0	0	0
CAT&CHEM INV.	0	0	0	70	0	0	0	0	0	0
SPARE&WAREHOUS	0	0	0	9500	9500	0	0	0	0	0
CASH ON HAND	0	0	0	0	11110	0	0	0	0	0
A/C REC.-PAY.	0	0	0	0	55680	0	0	0	0	0
REFUND OF LOANS	0	0	0	0	0	0	0	0	0	0
EQUITY	0	0	0	0	0	0	0	0	0	0
DIVIDEND	0	0	0	0	0	0	0	0	0	0
TOTAL	0	189990	474970	327930	104550	0	0	0	0	0
* CASH IN										
RETAINED EARN.	0	0	0	0	163051	330099	330099	330099	330099	330099
PROFIT AFT TAX	0	0	0	0	104388	271436	271436	271436	271436	271436
DEPRECIATION	0	0	0	0	58663	58663	58663	58663	58663	58663
LOANS	0	189990	474970	327930	0	0	0	0	0	0
EQUITY	0	189970	474970	327930	0	0	0	0	0	0
CAPITAL	0	0	0	0	0	0	0	0	0	0
TOTAL	0	189990	474970	327930	163051	330099	330099	330099	330099	330099
* SURPLUS	0	0	0	0	58501	330099	330099	330099	330099	330099
* CUM. SURPLUS	0	0	0	0	58501	388600	718699	1048798	1378897	1708996
* BALANCE OF LOANS	0	189990	664960	992890	992890	992890	992890	992890	992890	992890
EQUITY	0	189990	664960	992890	992890	992890	992890	992890	992890	992890

- Continued -

F U N D S   O U T L O O K

10**030OLL — E501

	11 1994	12 1995	13 1996	14 1997	15 1998	16 1999	17 2000	18 2001	19 2002	20 2003
* CASH OUT										
INVESTMENT	0	0	0	0	0	0	0	0	0	0
CONST. COSTS	0	0	0	0	0	0	0	0	0	0
PAIDUP ROYALTY	0	0	0	0	0	0	0	0	0	0
INIT.CAT&CHEM.	0	0	0	0	0	0	0	0	0	0
PREOPE. EXPNS.	0	0	0	0	0	0	0	0	0	0
OIL INV.	0	0	0	0	0	0	0	0	0	0
CAT&CHEM INV.	0	0	0	0	0	0	0	0	0	0
SPARE&WAREHOUS	0	0	0	0	0	0	0	0	0	0
CASH ON HAND	0	0	0	0	0	0	0	0	0	0
A/C REC.-PAY.	0	0	0	0	0	0	0	0	0	0
REFUND OF LOANS	0	0	0	0	0	0	0	0	0	0
EQUITY	0	0	0	0	0	0	0	0	0	0
DIVIDEND	0	0	0	0	0	0	0	0	0	0
TOTAL	0	0	0	0	0	0	0	0	0	0
* CASH IN										
RETAINED EARN.	330099	330099	330099	330099	330099	330099	330099	330099	330099	330099
PROFIT AFT TAX	271436	271436	271436	271436	271436	271436	271436	271436	271436	271436
DEPRECIATION	58663	58663	58663	58663	58663	58663	58663	58663	58663	58663
LOANS	0	0	0	0	0	0	0	0	0	0
EQUITY	0	0	0	0	0	0	0	0	0	0
CAPITAL	0	0	0	0	0	0	0	0	0	0
TOTAL	330099	330099	330099	330099	330099	330099	330099	330099	330099	330099
* SURPLUS	330099	330099	330099	330099	330099	330099	330099	330099	330099	330099
* CUM. SURPLUS	2039095	2369194	2699293	3029392	3359491	3689590	4019689	4349788	4679887	5009986
* BALANCE OF LOANS	992890	992890	992890	992890	992890	992890	992890	992890	992890	992890
EQUITY	992890	992890	992890	992890	992890	992890	992890	992890	992890	992890

- Continued -

E501

FUNDS OUTLOOK

10**03DHLL

	21 2004	22 2005	23 2006	24 2007	SUM
* CASH OUT					
INVESTMENT	0	0	0	0	1097440
CONST. COSTS	0	0	0	0	949940
PAIDUP ROYALTY	0	0	0	0	2130
INIT.CAT&CHEM.	0	0	0	0	5100
PREOPE. EXPNS.	0	0	0	0	16640
OIL INV.	0	0	0	0	37770
CAT&CHEM INV.	0	0	0	0	70
SPARE&WAREHOUS	0	0	0	0	19000
CASH ON HAND	0	0	0	0	11110
A/C REC.-PAY.	0	0	0	0	55680
REFUND OF LOANS	0	0	0	0	0
EQUITY	0	0	0	0	0
DIVIDEND	0	0	0	0	0
TOTAL	0	0	0	0	1097440
* CASH IN					
RETAINED EARN.	318367	300768	300768	300768	6335203
PROFIT AFT TAX	283169	300768	300768	300768	5361397
DEPRECIATION	35198	0	0	0	973809
LOANS	0	0	0	0	992890
EQUITY	0	0	0	0	992890
CAPITAL	0	0	0	0	0
TOTAL	318367	300768	300768	300768	7328092
* SURPLUS	318367	300768	300768	300768	6230653
* CUM. SURPLUS	5328352	5629119	5929886	6230653	
* BALANCE OF LOANS	992890	992890	992890	992890	
EQUITY	992890	992890	992890	992890	

- Continued -

CASH FLOW ANALYSIS

10**03DOLL 6501

YEAR	INVEST-MENT (-)	PROFIT AFTER TAX (+)	DEPREC-IATION (+)	CASH FLOW	D.C.F.	DISCOUNT RATE
1984	0	0	0	0	0	1.0000
1985	189990	0	0	-189990	-154658	0.8140
1986	474970	0	0	-474970	-314739	0.6626
1987	327930	0	0	-327930	-176892	0.5394
1988	104550	104388	58663	58501	25688	0.4391
1989	0	271436	58663	330099	117993	0.3574
1990	0	271436	58663	330099	96050	0.2910
1991	0	271436	58663	330099	78188	0.2369
1992	0	271436	58663	330099	63647	0.1928
1993	0	271436	58663	330099	51811	0.1570
1994	0	271436	58663	330099	42176	0.1278
1995	0	271436	58663	330099	34333	0.1040
1996	0	271436	58663	330099	27948	0.0847
1997	0	271436	58663	330099	22750	0.0689
1998	0	271436	58663	330099	18520	0.0561
1999	0	271436	58663	330099	15076	0.0457
2000	0	271436	58663	330099	12272	0.0372
2001	0	271436	58663	330099	9990	0.0303
2002	0	271436	58663	330099	8132	0.0246
2003	0	271436	58663	330099	6620	0.0201
2004	0	283169	35198	318367	5197	0.0163
2005	0	300768	0	300768	3997	0.0133
2006	0	300768	0	300768	3254	0.0108
2007	0	300768	0	300768	2648	0.0088
SUM	1097440	5361397	973809	5237764	0	

RATE OF RETURN 22.85 %

PAYOUT PERIOD 3.46 YEAR

ORINOCO HVY-O PJ

ITEM NO. BASEBASE
CASE NO. M501

E C O N O M I C

E V A L U A T I O N

P R O G R A M

JOB NO.
DATE 19800722
ENGINEER T.SHINDO

P R O F I T & L O S S

10**03DOLL M501

	1	2	3	4	5	6	7	8	9	10
	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993
* REVENUE										
IMP. CRUDE	0	0	0	0	625763	1251525	1251525	1251525	1251525	1251525
ELECTRIC POWER	0	0	0	0	12139	24278	24278	24278	24278	24278
FUEL GAS	0	0	0	0	0	0	0	0	0	0
INT. RECEIVED	0	0	0	0	0	0	0	0	0	0
TOTAL	0	0	0	0	637902	1275803	1275803	1275803	1275803	1275803
* EXPENSES										
RAW MATERIAL	0	0	0	0	263555	527110	527110	527110	527110	527110
CAT.&CHEM.	0	0	0	0	22680	22680	22680	22680	22680	22680
UTILITIES	0	0	0	0	0	0	0	0	0	0
OPERAT'G SUPPLY	0	0	0	0	1515	1515	1515	1515	1515	1515
TAXES	0	0	0	0	0	0	0	0	0	0
INSURANCES	0	0	0	0	0	0	0	0	0	0
MAINTE.& REPAIR	0	0	0	0	33240	33240	33240	33240	33240	33240
OPERATING LABOR	0	0	0	0	22572	22572	22572	22572	22572	22572
PLANT OH	0	0	0	0	9932	9932	9932	9932	9932	9932
DEPRECIATION	0	0	0	0	63498	63498	63498	63498	63498	63498
INT. PAID	0	0	0	0	0	0	0	0	0	0
TOTAL	0	0	0	0	416991	680546	680546	680546	680546	680546
* PROFIT BEF. TAX	0	0	0	0	220911	595257	595257	595257	595257	595257
* INCOME TAX	0	0	0	0	110455	297628	297628	297628	297628	297628
* PROFIT AFT. TAX	0	0	0	0	110455	297629	297629	297629	297629	297629
* CUM. PROFIT	0	0	0	0	110455	408084	705713	1003341	1300969	1598597

- Continued -

P R O F I T & L O S S

10**03DOLL M501

	11 1994	12 1995	13 1996	14 1997	15 1998	16 1999	17 2000	18 2001	19 2002	20 2003
* REVENUE										
IMP. CRUDE	1251525	1251525	1251525	1251525	1251525	1251525	1251525	1251525	1251525	1251525
ELECTRIC POWER	24278	24278	24278	24278	24278	24278	24278	24278	24278	24278
FUEL GAS	0	0	0	0	0	0	0	0	0	0
INT. RECEIVED	0	0	0	0	0	0	0	0	0	0
TOTAL	1275803	1275803	1275803	1275803	1275803	1275803	1275803	1275803	1275803	1275803
* EXPENSES										
RAW MATERIAL	527110	527110	527110	527110	527110	527110	527110	527110	527110	527110
CAT.&CHEM.	22680	22680	22680	22680	22680	22680	22680	22680	22680	22680
UTILITIES	0	0	0	0	0	0	0	0	0	0
OPERAT'G SUPPLY	1515	1515	1515	1515	1515	1515	1515	1515	1515	1515
TAXES	0	0	0	0	0	0	0	0	0	0
INSURANCES	0	0	0	0	0	0	0	0	0	0
MAINTE.& REPAIR	33240	33240	33240	33240	33240	33240	33240	33240	33240	33240
OPERATING LABOR	22572	22572	22572	22572	22572	22572	22572	22572	22572	22572
PLANT OH	9932	9932	9932	9932	9932	9932	9932	9932	9932	9932
DEPRECIATION	63498	63498	63498	63498	63498	63498	63498	63498	63498	63498
INT. PAID	0	0	0	0	0	0	0	0	0	0
TOTAL	680546	680546	680546	680546	680546	680546	680546	680546	680546	680546
* PROFIT BEF. TAX	595257	595257	595257	595257	595257	595257	595257	595257	595257	595257
* INCOME TAX	297628	297628	297628	297628	297628	297628	297628	297628	297628	297628
* PROFIT AFT. TAX	297629	297629	297629	297629	297629	297629	297629	297629	297629	297629
* CUM. PROFIT	1896225	2193853	2491481	2789109	3086737	3384365	3681993	3979621	4277249	4574877

- Continued -

P R O F I T & L O S S

10**030OLL M501

	21	22	23	24	
	2004	2005	2006	2007	SUM
* REVENUE					
IMP. CRUDE	1251525	1251525	1251525	1251525	24404704
ELECTRIC POWER	24278	24278	24278	24278	473427
FUEL GAS	0	0	0	0	0
INT. RECEIVED	0	0	0	0	0
TOTAL	1275803	1275803	1275803	1275803	24878080
* EXPENSES					
RAW MATERIAL	527110	527110	527110	527110	10278623
CAT.&CHEM.	22680	22680	22680	22680	453600
UTILITIES	0	0	0	0	0
OPERAT'G SUPPLY	1515	1515	1515	1515	30295
TAXES	0	0	0	0	0
INSURANCES	0	0	0	0	0
MAINTE.& REPAIR	33240	33240	33240	33240	664800
OPERATING LABOR	22572	22572	22572	22572	451440
PLANT OH	9932	9932	9932	9932	198634
DEPRECIATION	38099	0	0	0	1054068
INT. PAID	0	0	0	0	0
TOTAL	655147	617048	617048	617048	13131471
* PROFIT BEF. TAX	620656	658755	658755	658755	11746673
* INCOME TAX	310328	329377	329377	329377	5873335
* PROFIT AFT. TAX	310328	329378	329378	329378	5873336
* CUM. PROFIT	4885205	5214582	5543959	5873336	

– Continued –

F U N D S   O U T L O O K

10##03DOLL M501

	1 1984	2 1985	3 1986	4 1987	5 1988	6 1989	7 1990	8 1991	9 1992	10 1993
* CASH OUT										
INVESTMENT	0	201960	504890	366120	115190	0	0	0	0	0
CONST. COSTS	0	201960	504890	302730	0	0	0	0	0	0
PAIDUP ROYALTY	0	0	0	1200	1190	0	0	0	0	0
INIT.CAT&CHEM.	0	0	0	25260	0	0	0	0	0	0
PRODPE. EXPNS.	0	0	0	8320	8320	0	0	0	0	0
OIL INV.	0	0	0	18210	18200	0	0	0	0	0
CAT&CHEM INV.	0	0	0	100	0	0	0	0	0	0
SPARE&WAREHOUS	0	0	0	10100	10100	0	0	0	0	0
CASH ON HAND	0	0	0	0	14990	0	0	0	0	0
A/C REC.-PAY.	0	0	0	0	62390	0	0	0	0	0
REFUND OF LOANS	0	0	0	0	0	0	0	0	0	0
EQUITY	0	0	0	0	0	0	0	0	0	0
DIVIDEND	0	0	0	0	0	0	0	0	0	0
TOTAL	0	201960	504890	366120	115190	0	0	0	0	0
* CASH IN										
RETAINED EARN.	0	0	0	0	173954	361127	361127	361127	361127	361127
PROFIT AFT TAX	0	0	0	0	110455	297629	297629	297629	297629	297629
DEPRECIATION	0	0	0	0	63498	63498	63498	63498	63498	63498
LOANS	0	201960	504890	366120	0	0	0	0	0	0
EQUITY	0	201960	504890	366120	0	0	0	0	0	0
CAPITAL	0	0	0	0	0	0	0	0	0	0
TOTAL	0	201960	504890	366120	173954	361127	361127	361127	361127	361127
* SURPLUS	0	0	0	0	58764	361127	361127	361127	361127	361127
* CUM. SURPLUS	0	0	0	0	58764	419890	781017	1142143	1503269	1864395
* BALANCE OF LOANS	0	201960	706850	1072969	1072969	1072969	1072969	1072969	1072969	1072969
EQUITY	0	201960	706850	1072969	1072969	1072969	1072969	1072969	1072969	1072969

- Continued -

F U N D S   O U T L O O K

10**03DOLL

M501

	11	12	13	14	15	16	17	18	19	20
	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003
* CASH OUT										
INVESTMENT	0	0	0	0	0	0	0	0	0	0
CONST. COSTS	0	0	0	0	0	0	0	0	0	0
PAIDUP ROYALTY	0	0	0	0	0	0	0	0	0	0
INIT.CAT&CHEM.	0	0	0	0	0	0	0	0	0	0
PREOPE. EXPNS.	0	0	0	0	0	0	0	0	0	0
OIL INV.	0	0	0	0	0	0	0	0	0	0
CAT&CHEM INV.	0	0	0	0	0	0	0	0	0	0
SPARE&WAREHOUS	0	0	0	0	0	0	0	0	0	0
CASH ON HAND	0	0	0	0	0	0	0	0	0	0
A/C REC.-PAY.	0	0	0	0	0	0	0	0	0	0
REFUND OF LOANS	0	0	0	0	0	0	0	0	0	0
EQUITY	0	0	0	0	0	0	0	0	0	0
DIVIDEND	0	0	0	0	0	0	0	0	0	0
TOTAL	0	0	0	0	0	0	0	0	0	0
* CASH IN										
RETAINED EARN.	361127	361127	361127	361127	361127	361127	361127	361127	361127	361127
PROFIT AFT TAX	297629	297629	297629	297629	297629	297629	297629	297629	297629	297629
DEPRECIATION	63498	63498	63498	63498	63498	63498	63498	63498	63498	63498
LOANS	0	0	0	0	0	0	0	0	0	0
EQUITY	0	0	0	0	0	0	0	0	0	0
CAPITAL	0	0	0	0	0	0	0	0	0	0
TOTAL	361127	361127	361127	361127	361127	361127	361127	361127	361127	361127
* SURPLUS	361127	361127	361127	361127	361127	361127	361127	361127	361127	361127
* CUM. SURPLUS	2225521	2586647	2947773	3308899	3670025	4031151	4392277	4753403	5114529	5475655
* BALANCE OF LOANS	1072969	1072969	1072969	1072969	1072969	1072969	1072969	1072969	1072969	1072969
EQUITY	1072969	1072969	1072969	1072969	1072969	1072969	1072969	1072969	1072969	1072969

- Continued -

F U N D S   O U T L O O K

10**03DOLL

M501

	21 2004	22 2005	23 2006	24 2007	SUM
* CASH OUT					
INVESTMENT	0	0	0	0	1188160
CONST. COSTS	0	0	0	0	1009780
PAIDUP ROYALTY	0	0	0	0	2390
INIT.CAT&CHEM.	0	0	0	0	25260
PRLOPE. EXPNS.	0	0	0	0	16640
OIL INV.	0	0	0	0	36410
CAT&CHEM INV.	0	0	0	0	100
SPARE&WAREHOUS	0	0	0	0	20200
CASH ON HAND	0	0	0	0	14990
A/C REC.-PAY.	0	0	0	0	62390
REFUND OF LOANS	0	0	0	0	0
EQUITY	0	0	0	0	0
DIVIDEND	0	0	0	0	0
TOTAL	0	0	0	0	1188160
* CASH IN					
RETAINED EARN.	348427	329378	329378	329378	6927403
PROFIT AFT TAX	310328	329378	329378	329378	5873336
DEPRECIATION	38099	0	0	0	1054068
LOANS	0	0	0	0	1072969
EQUITY	0	0	0	0	1072969
CAPITAL	0	0	0	0	0
TOTAL	348427	329378	329378	329378	8000370
* SURPLUS	348427	329378	329378	329378	6812213
* CUM. SURPLUS	5824082	6153459	6482836	6812213	
* BALANCE OF LOANS	1072969	1072969	1072969	1072969	
EQUITY	1072969	1072969	1072969	1072969	

- Continued -

C A S H   F L O W   A N A L Y S I S

10**03DOLL. M501

YEAR	INVEST- -MENT (-)	PROFIT AFTER TAX (+)	DEPREC- -IATION (+)	CASH FLOW	D.C.F.	DISCOUNT RATE
1984	0	0	0	0	0	1.0000
1985	201960	0	0	-201960	-164113	0.8126
1986	504890	0	0	-504890	-333389	0.6603
1987	366120	0	0	-366120	-196452	0.5366
1988	115190	110455	63498	58764	25622	0.4360
1989	0	297629	63498	361127	127952	0.3543
1990	0	297629	63498	361127	103974	0.2879
1991	0	297629	63498	361127	84489	0.2340
1992	0	297629	63498	361127	68656	0.1901
1993	0	297629	63498	361127	55790	0.1545
1994	0	297629	63498	361127	45335	0.1255
1995	0	297629	63498	361127	36839	0.1020
1996	0	297629	63498	361127	29936	0.0829
1997	0	297629	63498	361127	24326	0.0674
1998	0	297629	63498	361127	19767	0.0547
1999	0	297629	63498	361127	16063	0.0445
2000	0	297629	63498	361127	13053	0.0361
2001	0	297629	63498	361127	10607	0.0294
2002	0	297629	63498	361127	8619	0.0239
2003	0	297629	63498	361127	7004	0.0194
2004	0	310328	38099	348427	5491	0.0158
2005	0	329378	0	329378	4218	0.0128
2006	0	329378	0	329378	3428	0.0104
2007	0	329378	0	329378	2785	0.0085
SUM	1188160	5873336	1054068	5739245	-1	

RATE OF RETURN 23.06 %

PAYOUT PERIOD 3.43 YEAR