THE REPUBLIC OF IRAQ

TECHNICAL STUDY REPORT
ON
EXPORT OIL REFINERY

JANUARY 1977

JAPAN INTERNATIONAL COOPERATION AGENCY

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M·P I

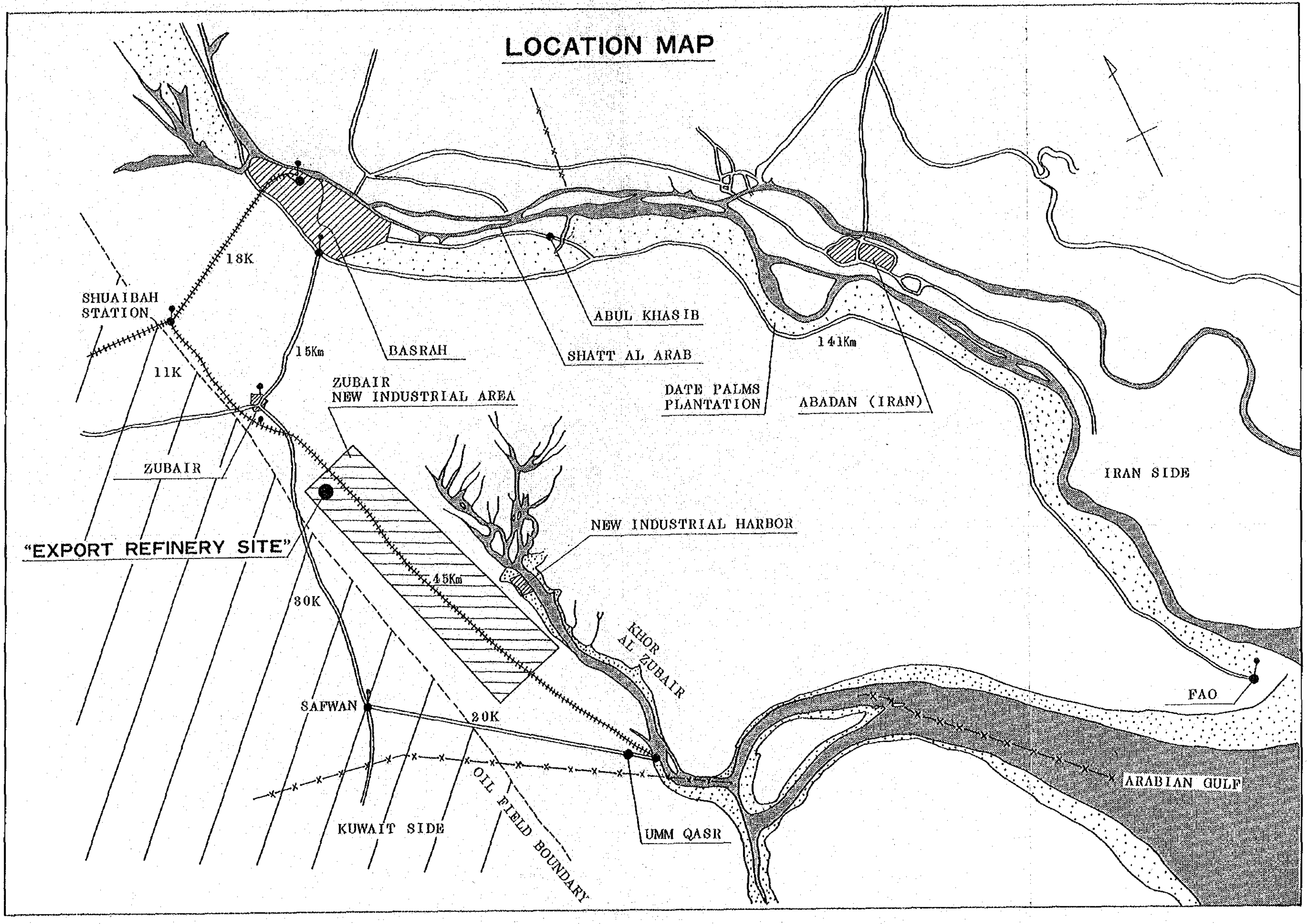
PREFACE

In accordance with "the Agreement on Economic and Technical Cooperation between the Government of Japan and the Government of the Republic of Iraq", our government entrusted the Japan International Cooperation Agency with the task of technical survey for developing the export oil refinery scheduled in Iraq. A survey team organized by the agency was dispatched for field studies in connection with the export oil refinery project. After their return to Japan, technical information gathered in the survey was used for the design work as well as for completing a report. Technical studies involved in this report are for a refinery planning, basic planning and basic design of each process unit and preparation of a basic design for off-site utilities, etc.

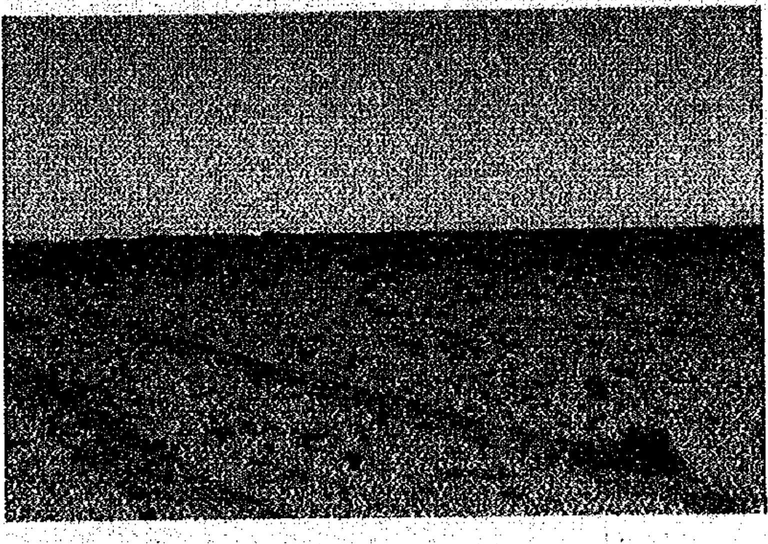
In presenting this report, we hope that it will contribute towards the promotion of the export oil refinery project and its associated industrial development, thus increasing the econimic growth of the Republic of Iraq.
Through this, we also hope to strengthen the friendly relationship between Japan and the Republic of Iraq.

January, 1977

Shinsaku HOGEN
President
Japan International Cooperation Agency

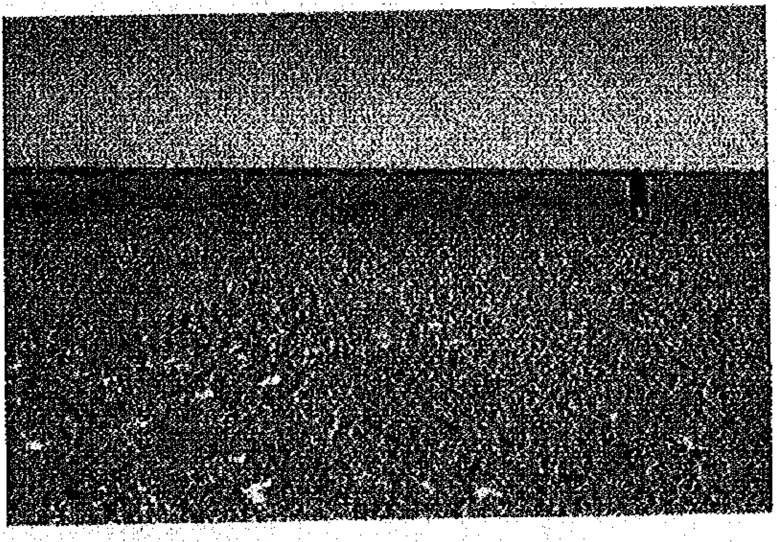


LOCATION MAP



Refinery Job Site

Product Pipeline Route
at Shatt Al Basrah Canal



Effluent Pipeline Route through
Flooded Area near Khor Al Zubair

CONTENTS

PAGE

DRAWING INDEX

Chapter 1.

Introduction

Chapter 1. Introduction

1.1 Survey

1.1.1 General

The present survey was carried out in accordance with the instructions of Japan International Cooperation Agency by a survey group which was organized in Japan in July, 1976 and despatched to Republic of Iraq from July 7 to July 16, 1976. The results of the survey were applied to preparing this report in Japan as basic data after the group returned to Japan.

1.1.2 Purposes of survey

The survey was for the purpose of collecting technical information required for executing the export refinery project being planned by the State Company of Oil Project of Republic of Iraq. The outline of the survey was listed up as follows:

(a) Surveying the anticipated refinery construction site, which is Zubair, south of Basrah, to apply to the overall refinery plan which will be the fundamental of the present project.

(b) Study on the pipeline routes under plan together with studying the design basis required for each route.

(c) Survey on the planned water intake point together with examination of water quality, etc.

(d) Survey on the electric power supply condition for establishing fundamental project plan.

(e) Collection of other necessary information

1.1.3 Organization of survey group

The group was organized with the following eight specialists of Japan Consulting Institute.

(a) Chief	Period of Survey
Yoshitatsu Iwamoto (General)	July 7, 1976 - July 13, 1976
(b) Surveyors	
Makoto Tanuma (Project)	" "
Toshio Ando (Civil and Architectural)	" "
Akio Toyama (Project)	" "
Tsutomu Kishi (Construction)	" "
Yuji Yoshii (Construction)	" "
Kunio Yanagizawa (Pipeline)	July 7, 1976 - July 16, 1976
Akio Otani (Pipeline)	" "

Members of the survey group wish to take this opportunity to express their gratitude to the State Company of Oil Project of Republic of Iraq for its cooperation and assistance granted to the group in executing the survey.

1.2 Summary of report

1.2.1 Introduction

This technical survey report has been prepared basically according to the TENDER DOCUMENTS for the EXPORT REFINERY PROJECT furnished by the STATE COMPANY OF OIL PROJECT (hereinafter call SCOP).

This report explains the result of our studies concerning the crude assay and refinery scheme from SCOP's assumed standpoint. And this report explains our basic thought and conditions concerning the related process units, utility facilities, off-site tankage facilities and pipeline systems for this refinery project, and also explains the basic conditions for execution of design engineering from SCOP's assumed standpoint for these design items.

Especially refinery layout plan is studied as a modern refinery according to the drawing which are attached to the TENDER DOCUMENTS.

Namely, some deviations from the TENDER DOCUMENTS and some clarifications of the TENDER DOCUMENTS are explained in this report.

1.2.2 Summary of each chapter

(1) Summary of Chapter 2. Crude Assay

In proceeding with the process design on the basis of using N-Rumaila and Mischrif crudes as feedstocks can be determined by referring to the crude assay given in the TENDER DOCUMENTS Vol. VI d.

However, regarding the N-Rumaila crude, the TENDER DOCUMENTS require that recalculations and revisions shall be made based on the TBP curve (Fig. 9.1, Para. 15, Sect. 9, Vol. II), probably due to obsolescence of the present N-Rumaila crude assay.

Therefore, detailed procedures concerning the work of revising the N-Rumaila crude assay are given herein and the result of our reviews, namely the values and tables obtained as a result of the revisions and reviews, are attached.

The given crude assay for the Mischrif crude is not sufficient for design purposes.

However, using this crude assay as a basis and supplementing it with the results of our studies, products that can be obtained from this Mischrif crude and the results of our calculations on the presumed yields of products and on the presumed yields of products and their characteristics are shown in Item 4.2, CRUDE UNIT.

(2) Summary of Chapter 3. Refinery Shceme

We have carefully studied and revised the refinery scheme (as set in the TENDER DOCUMENTS) in view of the overall revaluation and revisions necessitated by the change in feed-stock from Buzurgan and Abu-Ghrab crude to Mischrif crude.

The revisions have been made based on the crude assay mentioned the preceding section.

The relative schematic drawing is attached as Dwg. No. 00-G-001.

That is, the refinery scheme is composed of groups of PROCESS UNITS comprising trains for treating N-Rumaila crude and Mischrif crude respectivery.

7,500,000 T/Y of crude can be treated in the respective trains.

The capacities, products and quantities for the units and related common facilities of each train have been established as a result of studies and revisions made with full consideration to the philosophy of the TENDER DOCUMENTS and the basis of the revised refinery scheme is discussed.

(3) Summary of Chapter 4. Process Design

(a) General

This section describes the basic philosophy on PROCESS UNIT design and, at the same time, explains the reasons for selecting the processes from the viewpoint of the respective PROCESS UNITS.

Namely, the atmospheric distillation residue direct desulfurization unit of GULF Co. and paraxylene extraction unit of UOP (using naphtha as feed-stock) are designated in the TENDER DOCUMENTS.

However, the other units are to be reviewed and selected by us.

Therefore it has been decided to also adopt the licenced processes of UOP, UHDE and TOPSØE.

The results of our studies on the treatment capacities (design capacity) of the respective PROCESS UNITS based on the refinery scheme mentioned the preceding section are given in the form of a table.

These treatment capacities have been decided based on the capacities shown in the aforementioned refinery scheme, taking operating efficiencies etc. into account.

4.2 - 4.12 CRUDE DISTILLATION UNIT and other PROCESS UNITS.

In Para. 4.3 through 4.12 the details concerning the respective PROCESS UNITS are described as follows.

(i) Process design basis for RESPECTIVE UNITS

In general, the capacities of the UNITS, properties of feed-stock, product specifications and process design outlines, which constitute the basis of the process design, are described herein.

(ii) Explanations on respective PROCESS UNITS

The types of equipment comprising each PROCESS UNITS and the related flow are explained with the respective flow sheets attached.

(iii) Characteristics of respective PROCESS UNITS

It is emphasized and explained that the respective unit are designed to possess economical and functional characteristics.

(iv) Expected yields and product properties

These will serve as bases for establishing guarantee values for the RESPECTIVE UNIT in future.
Though not especially specified as guarantee values, these are important items which also constitute factors in fixing the refinery scheme.

(v) Tie-in conditions of each UNIT

Unless the tie-in conditions between each UNIT are mutually confirmed, the design and operation of the UNITS will be greatly hindered.
Therefore, these are clearly given as important data that will be necessary for the overall restudy of the refinery.

(4) Summary of Chapter 5. Refinery Utility Design

The refinery utility facilities comprise the water supply facility, steam facility, instrumentation and plant air facility, fuel facility, inert gas facility, etc.
The design philosophy for each facility which was reviewed and established by us with the consideration to the design policy in the TENDER DOCUMENTS is described, and the basic data for deciding the major equipment are shown.

Schematic system drawings of the respective facilities are attached to assist the explanations.
The capacities of the major utility facilities are as fallows.

Filtered water tank : 80,000 m^3 x 1 ea
PURE WATER UNIT : 840 T/H x 3 ea
COOLING WATER UNIT : 25,000 T/H x 2 ea
Boiler : 135 T/H x 4 ea
Air compressor : 13,000 Nm^3/hr x 2 ea
NITROGEN GENERATING UNIT: 400 Nm^3/hr x 3 ea

As the steam levels inside the refinery are not given in the TENDER DOCUMENTS, we have reviewed and stipulated them as follows.

High pressure steam : 32 kg/cm^2G 350°C
Medium pressure steam : 15 kg/cm^2G 350°C and saturation
Low pressure steam : 3.5 kg/cm^2G saturation

The interface, break point between the applicable ranges of temperature for water coolers and air coolers have been decided with consideration to economics and designability based on the proposed refinery site.

(5) Summary of Chapter 6. Refinery Off-site Tankage Facility

As the crude has been changed and the refinery scheme has been revised, the off-site tank facilities must also be adjusted accordingly.
However, we have decided the optimum design base according to the policy of respecting the requirements of the TENDER DOCUMENTS as far as possible.

Accordingly, deviations from the TENDER DOCUMENTS have been clarified and the design bases concerning the crude tank facilities, intermediate tank facilities, product components tank facilities, product tank facilities, tank ancillary facilities, etc. are described.

Moreover, a list of the related tanks is attached and the oil storage and handling capacities of the refinery are shown.
Further, as explained in Chapter 7, Refinery layout the tank layout attached to the TENDER DOCUMENTS would result in a basically unreasonable arrangement due to the necessity of expanding the plot for the PROCESS UNITS and associated facilities.
With consideration also to the results of studies on the oil dike capacities, the overall layout has been restudied and the revised plan is attached.

(6) Summary of Chapter 7. Refinery Layout

(a) As described in the preceding section the overall plot plan resulting from our review is attached as a preparable alternative to the refinery plot plan attached to the TENDER DOCUMENTS.
The topographic site survey drawing presented together with the TENDER DOCUMENTS differs in scale from the abovementioned refinery plot plan and their relation is not clear at all.
Therefore, we propose that Bench Mark "A" of this survey drawing be set up at the north-western and of the refinery plot plan and the locations of the roads, buildings, equipment, etc, in the refinery be expressed with reference to this Bench Mark "A".

(b) As the classification of the units into groups shown in the refinery plot plan attached to the TENDER DOCUMENTS will lead to fundamental unreasonableness, regarding the changing of the said refinery plot plan and changes due to the impossibility of designing an adequate layout unless the plots of the UNITS are enlarged, an overall PROCESS UNIT plot plan is attached, with explanations.

(c) Also regarding the plot for the utility facilities, in the same manner, as an extension of the area and changes in facility locations are required under the plot plan attached to the TENDER DOCUMENTS, a drawing is attached, with explanations.

(d) One of the items relating to the refinery plot plant is the ground elevation of the respective facilities. However, this matter also is not definitely indicated in the TENDER DOCUMENTS.
Therefore, we have decided the elevations as a result of field investigations based on the topographic site survey drawings and have shown them in the attached drawing.

(7) Summary of Chapter 8. Pipeline Design

Of the pipeline, the location of the FAO terminal has not yet been decided.
Therefore, excluding the pipelines related to the above, we have only reviewed the lines related to this refinery and have proposed the respective routes as per the drawings attached.
The design bases, requirements and standards to be applied to the pipeline are described.
Regarding the respective pipelines, the design flow rates reviewed and decided by us, the pipe size adopted based on such flow rates, loading pump facility specifications, etc. are explained with reference to flow diagrams.

Especially, regarding the raw water supply line of the utility facilities, a route that is markedly different from that designated in the TENDER DOCUMENTS is proposed as a result of various inconveniences found with regard to the supply points during site investigations.
The old refinery site is proposed as the location for the MUFTIEH for loading benzene and paraxylene moreover, the drain water pipeline will be installed on racks as it will be routed through swampy areas.

(8) Summarly of Chapter 9. Design Engineering

In proceeding with the design of the respective PROCESS UNITS, utility facilities, off-site tank facilities, etc. it is necessary to perform the design work of the respective design categories based on a unified policy. The bases are confirmed and explained in this section.

(a) Basic engineering design data

These clauses concerning the design bases stipulate general fundamental matters not included in the above-mentioned respective design categories and mention the basic philosophy concerning the laws and regulations, standards and standard specifications to be applied. Mareover, the basic philosophy on such matters as the design basis for utilities, conditions of locationing, measures against environmental pollution, etc., are explained and confirmed.
As the power supply capacity (64 MVA) stipulated in the TENDER DOCUMENTS can not meet the power requirements of this vast refinery, we have stipulated the maximum required incoming capacity of the electrical facilities based on the results of our calculations.
Especially, the properties of the raw water for the water supply as indicated in the TENDER DOCUMENTS refer to analysis values of the water at one

point at a certain time and does not cover annual fluctuations, changes of the values with time etc.
Therefore, it would be dangerous to adopt such values as basic design data for such a gigantic project.
Accordingly, we have established design criteria for water quality based on the results of various studies.

(b) Equipment design

Important matters that should be especially considered and confirmed regarding the design bases of equipment such as furnaces, heat exchangers, air coolers, unfired pressure vessels, pumps and drivers, compressors and drivers, are raised and clarified.
That is, even in the case where the standard specifications given in the TENDER DOCUMENTS are to be followed, in principle, the special conditions in the case of exports from Japan in order to perform the design and manufacturing effectively and economically are also considered and explained.
Morever, regarding rotary equipment such as pumps and compressors, it is emphasized that in selecting manufacturers, equipment adequate to the process units are to be selected with care, upon adequate review.

(c) Piping design

It is mentioned, in principle, to follow the piping specification Class "M" indicated in the TENDER DOCUMENTS, and the basic philosophy for the piping is described with regard to the overall design procedures.
Further, the piping classes selected and adjusted by us are listed up and the purposes and scopes of use of the respective classes are clarified.

(d) Civil design

In the same manner as the preceding paragraph, the standard specifications included in the TENDER DOCUMENTS to be followed are taken up, and the basic design philosophy is mentioned.
Especially, regarding JIS Standards not included in the TENDER DOCUMENTS, the scope of their application, etc. are described.

(e) Steel structure design

(f) Building design

Both the abovementioned clauses are described in accordance with the same explanation summary as for (d).

(g) Electricals design

The specifications concerning the design, manufacture and inspection of the equipment and materials of the electrical facilities in this export refinery are described.
It is clearly mentioned that the Japanese Standards will be especially applied to electrical items to be procured in and delivered from Japan.
Moreover, the basic philosoly concerning the design of the electrical facilities in this refinery, power source, characteristics, allowable limits of capacity, etc. is described.

(h) Instrumentation design

Regarding the instrumentation design, the fact that the central aim of the design is safe and easy operation is described in this clause, and important changes with respect to the TENDER DOCUMENTS are described.

Further, regarding the adopted basic clauses of the detailed design of the instrumentation, their applied specifications are definitely clarified.

(i) Insulation and painting

In this clause, the basic philosophy concerning the prevention of burns such as personal protection together with the prevention of heat radiation or heat absorption such as by hot or cold insulation of the UNITS is described, and regarding painting and fireproofing, the scope of their application, procedures, etc. are stipulated.

(j) Design of fire fighting facilities

The outline of the design conforming to the TENDER DOCUMENTS is described and corrosion-proofing procedures for the underground piping are also included.

(k) Design of drainage system

The drainage systems will be designed in accordance with the specifications attached to the TENDER DOCUMENTS. Moreover, the basic design clauses for the underground drain piping, open ditches and rainfall intensity are confirmed.

1.3 Definition of terms

Definition of terms contained herein are mainly used for this technical study report.

(1) N-RUMAILA .... Rumaila north crude oil:

Nomenclatures of any process areas, facilities, etc. prefixed by this word mean relative trains. Assignment of Unit Nos. conforms to TENDER DOCUMENTS Vol. II Section 9, Para. 9.8.1.

(2) MISCHRIF .... Rumaila north mischrif zone crude oil:

Nomenclatures of any process areas, facilities, etc. prefixed by this word mean the relative trains. Assignment of Unit Nos. conforms to TENDER DOCUMENTS Vol. II Section 9, Para. 9.8.1

(3) CRUDE UNIT .... Crude distillation units

(4) N-HDS UNIT .... Naphtha desulfurization units

(5) K-HDS UNIT .... Kerosene desulfurization units

(6) GO-HDS UNIT .... Gas oil desulfurization unit

(7) CAT-REFORMER .... Catalytic reformer units

(8) GAS/AMINE UNIT ...... Gas recovery and amine treating unit

(9) SR-UNIT .... Sulfur recovery units Nos. Train A and Train B

(10) BTX-UNIT .... BTX extraction and fractionation unit

(11)	PAREX-UNIT	Paraxylene extraction unit
(12)	GULF HDS UNIT	Residue desulfurization units Train A and Train B
(13)	HYDROGEN UNIT	Hydrogen units
(14)	HDS COMPLEX	Processing units consisting of above (12) and (13) together with relative appurtenances.

Chapter 2.

Crude Assay

Chapter 2. Crude Assay

2.1 General

The process design is to be carried out on the basis of the N-RUMAILA and the MISCHRIF. The characteristics of the crudes to be employed for this design are given in TENDER DOCUMENTS Vol. VI-d, as crude assays.

The N-RUMAILA crude assay has been re-calculated and revised by us based on the TBP curve as indicated in Fig. 9.1 (TENDER DOCUMENTS Vol. II Sect. 9, page 15 of 33).

However, this revision work will not be guaranteed by us, it is understood that they will be confirmed by SCOP at the responsibility of SCOP.

Meanwhile, the MISCHRIF crude assay is retained in its original form.

The description of the revision work on the N-RUMAILA crude assay is mentioned in the next paragraph. The revised N-RUMAILA crude assay and the original MISCHRIF crude assay are attached.

2.2 Revision work on the N-Rumaila crude assay

(1) Introduction

The crude assay (on N-RUMAILA) prepared by IFP has been revised based on the TBP curve given in Fig. 9.1 (TENDER DOCUMENTS Vol. II Sect. 9, page 15 of 33) as follows:

(2) Method of revision

(a) Light hydrocarbons

C_4 and lighters are 2.5 vol% on crude, and C_5 and lighters are 5.4 vol% on crude. The composition of light hydrocarbon is assumed to be the same as that given in the assay prepared by IFP (hereinafter called the IFP assay).

	IFP Assay		Revised Assay	
	Vol% on Crude	wt% on Crude	Vol% on Crude	wt% on Crude
C_2	0.10	0.05	0.08	0.03
C_3	0.69	0.41	0.57	0.33
i-C_4	0.35	0.23	0.29	0.19
n-C_4	1.88	1.29	1.56	1.05
i-C_5	1.28	0.94	1.06	0.76
n-C_5	2.22	1.64	1.84	1.34
Total:	6.52	4.56	5.40	3.70

(b) Narrow cuts

The table given in Appendix I, page 14 of 62 of the IFP assay will be revised.

The specific gravities of narrow cut in the TBP cut range have been assumed to be correct.

The volume percentage of each cut is read from Fig. 9.1 (TENDER DOCUMENTS Vol. II Sect. 9, page 15 of 33), and shown on Table 4 given in Appendix II, page 4 of 18.

(c) Gasoline

The specific gravities of debutanized gasoline within the cut range are summarized based on the table prepared in the above Para. (b) and have been plotted to give a graph.

The sulfur content of each cut given in Appendix I, page 20 of 62 of the IFP assay has been assumed to be correct, and are summarized and plotted in the form of a graph.

The PONA analysis for the TBP cut range and the aromatic content shown at Appendix I, page 20 of 62, page 31 of 62 and page 42 of 62 of the IFP assay are assumed to be correct.

The other properties are regarded as being the same as those in the IFP assay.

(d) Kerosene

The kerosene cuts within a cut range of 50°C are summarized based on the table mentioned in the above Para. (b). The specific gravities of the cuts have been calculated and the TBP mid. points of the cuts graphed. The sulfur content will be checked by the same procedure based on Appendix I, page 20 of 62 of the IFP assay.

The other properties will be regarded as being the same as those in the IFP assay.

(e) Gas oil

The gas oil cuts within a cut range of 50 - 60°C are summarized and the specific gravities, sulfur content, pour point and aniline point of the cut calculated, and the TBP mid. points graphed.

The other properties are regarded as being the same as those in the IFP assay.

(f) Reduced crude

Since the above method (of summarizing narrow cuts) cannot be used, there is no alternative but to use the graph prepared by plotting the TBP cut points shown in Appendix I, page 53 of 62 of the IFP assay.

(3) Summary

The IFP assay has been revised by the following three methods:

(a) The volume percentages of the narrow cuts shown in Appendix I, page 14 of 62 of the IFP assay are revised based on Fig. 9.1 TENDER DOCUMENTS Vol. II Sect. 9, page 15 of 33 and the specific gravities of the TBP cut and the properties given in Appendix I, page 20 of 62 are regarded as correct.

(b) The properties estimated by summarizing those of the narrow cuts, using the table revised as mentioned in above (a), are used as a revised assay.

(c) The properties at the TBP cut points shown in the IFP assay are used.

The specific gravity curves, sulfur content, pour point and viscosity curves are obtained by method (A).

The specific gravities and sulfur content of gasoline, kerosene and gas oil and the pour point and aniline point of gas oil are obtained by method (B).

The properties of reduced crude and the PONA analysis of gasoline are those shown in the IFP assay.

Chapter 3.

Refinery Scheme

Chapter 3. Refinery Scheme

3.1 General

Since the feed stock has been changed to MISCHRIF, the refinery block flow indicated on Dwg. No. 52-E-1, 2 and 3 attached to the TENDER DOCUMENTS has been modified.

The modified refinery block flow diagram (Dwg. No. 00-G-001) is attached hereto.

The scheme has been planned on the basis that the refinery shall consist of two trains, each capable of processing 7,500,000 metric tons per annum (TPA) of crude oil and that each train shall be designed to process respectively MISCHRIF and N-RUMAILA. Each train will consist of a CRUDE UNIT, N-HDS UNIT, K-HDS UNIT, GULF HDS UNIT, and CAT-REFORMER.

The train to be used for MISCHRIF will consist of a GO-HDS UNIT in addition to the above types of units.

Also, it is understood that one GO-HDS UNIT will be provided in the future for the train used for N-RUMAILA.

The common units for both trains are a GAS/AMINE UNIT, HYDROGEN UNIT, SR-UNIT, BTX-UNIT and RAREX-UNIT.

The products to be produced in this refinery will be C_4 LPG, PETROCHEMICAL NAPHTHA, REGULAR and PREMIUM GASOLINE, BENZENE, P-XYLENE, KEROSENE, AVIATION TURBINE FUEL, AUTOMOTIVE DIESEL FUEL, HEATING OIL, 0.3 wt% SULFUR FUEL OIL, 1.0 wt% SULFUR FUEL OIL and SULFUR.

3.2 Basis

The following conditions have been considered in the modification of the scheme:

(1) The TBP cut ranges in crude distillation shall be the same for both trains.

(2) Maximum gasoline production shall be considered and CAT-REFORMER normal charge shall be TBP 65 - 175°C cut naphtha.

(3) TBP 150 - 175°C cut heavy naphtha shall not be produced.

(4) The production ratio of regular gasoline to premium gasoline shall be 87 : 13.

(5) The production ratio of automotive diesel fuel to heating oil shall be 43 : 57.

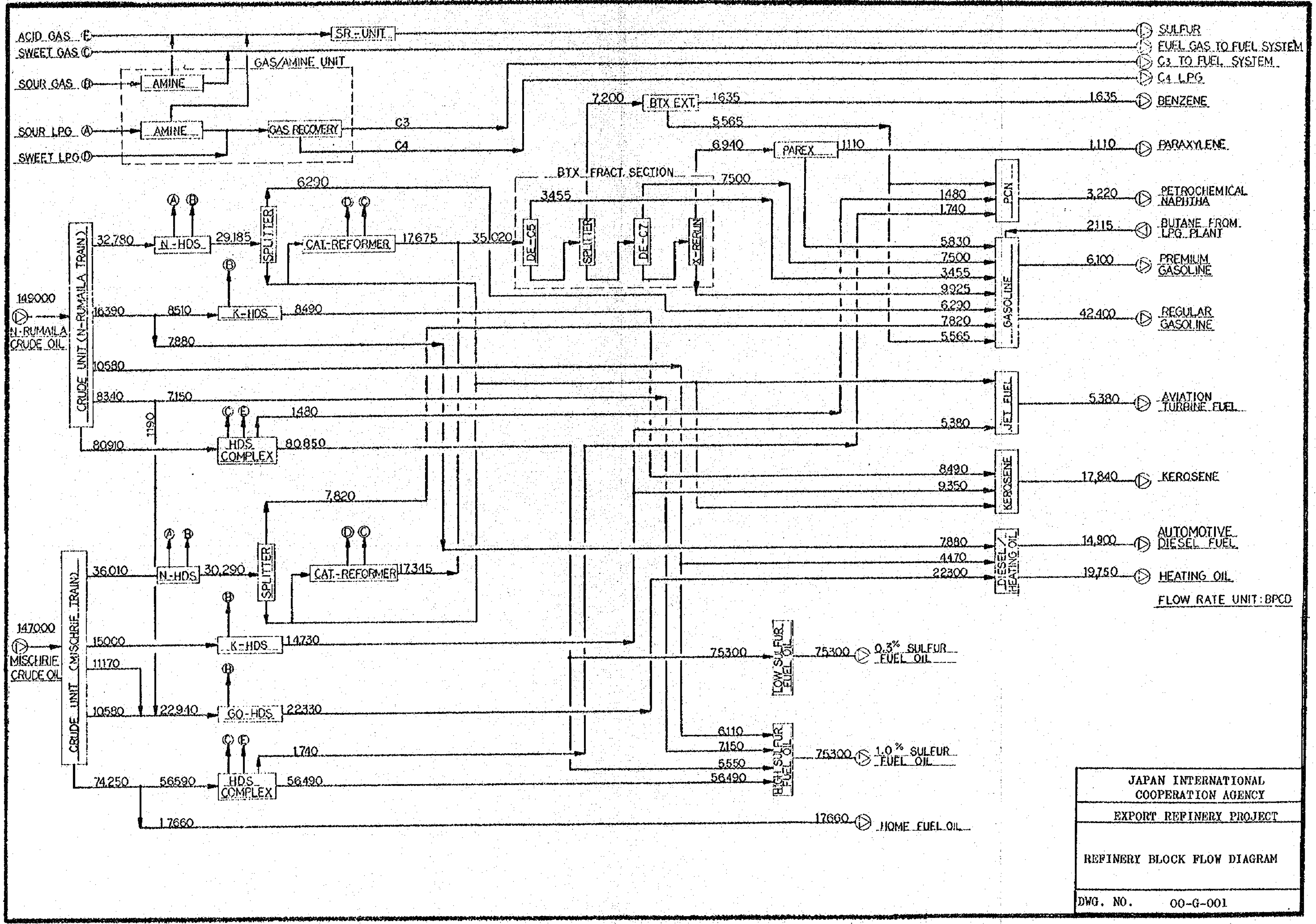
(6) Stabilized naphtha produced in GULF HDS UNIT shall be used as a component of petrochemical naphtha.

(7) The production rates of 1.0 wt% S fuel oil and 0.3 wt% S fuel oil shall be equal.

(8) The production rate of aviation turbine fuel shall be 250,000 TPA.

(9) The production rate of kerosene shall be left undecided.

(10) Benzene and p-xylene available in the CAT-REFORMER shall be recovered economically to a maximum extent.



JAPAN INTERNATIONAL COOPERATION AGENCY

EXPORT REFINERY PROJECT

REFINERY BLOCK FLOW DIAGRAM

DWG. NO. 00-G-001

Chapter 4.

Process Design

Chapter 4. Process Design

4.1 General

4.1.1 Process design philosphy

(1) Design parameters

To fix process flow diagram, generally the following case studies and optimization work are executed.

(i) Tower arrangements

How many side-cut streams of main tower are best?
How about proposing preflash tower?, etc.

(ii) Optimization of heat exchanger network

(iii) Concerning main tower, the best position of external reflux considering fractionation and tower diameter.

(iv) Optimum arrangement of air coolers and water coolers.

(v) Concerning heat exchangers, shell and tube side fluid selection considering heat transfer coefficients and maintenance.

(2) Process work

(a) Process engineerring calculations

(i) Physical properties

API data book

(ii) Process analysis and computer simulation

(iii) Unit operations

Equilibrium stage calculations (distillation, flash, absorbtion, etc)

(iv) Reaction kinetics

(v) Process stream separation and combination

(vi) Heat exchange, etc.

(b) Results of calculations

(i) Material balance

(ii) Heat and weight balance

(iii) Pressure drop survey

(iv) Rating summaries

Trays, exchangers, pumps and compressors, instruments line index, etc.

(c) Rating

(i) Columns

Trays:

Perforated tray, bubble tray, packing, etc.
- FRI method

Valve type tray, perform contact tray, etc.
- Proprietary supplier's method

Internal: Contracot's basic design manual

(ii) Drums

CONTRACTOR's basic design manual

(iii) Exchangers; Shell and tube exchangers; HTRI method
Air cooler; proprietary supplier's method

(iv) Pumps, compressors

Contractor's basic design manual

(v) Piping sizing, etc.

CONTRACTOR's basic design manual

(vi) Pressure relief systems

Contractor's basic design manual, etc.

4.1.2 Process selection

(1) AMINE UNIT

Regarding amine treating, process selection is amine solution selection. At present, the following types of amine are used for amine treating in refineries: MEA, DEA and ADIP.

Of these, DEA solution is used chiefly for gas treating in FCC unit where impurities such as COS and CS_2 are present. In case of the subject refinery consisting of no such units as FCC unit, either MEA or ADIP can be used.

In this case, the use of MEA has been considered for the following reasons:

(a) MEA treating process is not patented.

(b) Process operation is easy.

(c) Flexibility of the process unit is large.

(2) N-HDS, K-HDS and GO-HDS UNIT

All available naphtha, kerosene and gas oil processes are superior with outstanding features explained herein and non-disadvantageous without any demerits.

Therefore, any processes can be used, but UOP's processes have been selected. since we have had most experience in designing and constructing UOP's hydrodesulfurization units. If UOP's processes are decided on, we can proceed with all the work (from basic design to construction) smoothly.

Especially in case of UOP's processes, UOP will furnish basic design data on the reactor system only and we will provide all other basic design and detailed design.
This will add to flexibility in the design.

(3) CAT-REFORMER

In such cases as in this case where high octane gasoline and benzene and p-xylene are to be produced, very severe operation will be required for the CAT-REFORMER.

Meanwhile, H_2 rich gas available in the CAT-REFORMER will be used as make-up hydrogen gas in the distillate desulfurization units, and hydrogen for this use will be required to be of high purity and to be available at a constant rate. With consideration given to the above two points, UOP's continuous catalyst regeneration type platforming process has been selected.

(4) BTX UNIT

The Shell sulfolane process which was developed by Shell and is licensed by UOP, has been selected, with consideration given to the follwoing points:

(a) UOP's CCR platforming process has been selected for the CAT-REFORMER which will be upstream of this unit.

(b) UOP's PAREX process has been selected for the p-xylene extraction unit which will be downstream of this unit.

(c) UOP has had much experience in designing BTX extraction process units.

(d) Sulfolane to be used as extraction soluent is very excellent.

(e) We has had much experience in construction Sulfolane process units.

(5) SR UNIT

Sulfur recovery processes available for the acid gas from amine treating units and waste water treating units are generally limited to Clause type dry catalyst bed processes. Several Clause type sulfur recovery processes are available, but it is difficult to decide their superiority or inferiority by evaluation of their process characteristics, and hence many processes are very competitive with each other, as actual installations indicate.

To date, we have constructed numerous sulfur recovery units using Uhde's (Federal Republic of Germany) process, and have obtained quite satisfactory results. Therefore, in this case also, the Uhde process has been selected.

(6) GULF-HDS UNIT

Gulf Residual HDS Process has been selected in accordance with a designation in TENDER DOCUMENTS Vol. II Sect. 10, Page 2 of 33. We will provide a complemental process design based upon a Basic Process Design Specification furnished by Gulf Science and Technology Company.

(7) HYDROGEN UNIT

We has selected the licensors for hydrogen production unit, based on commercially proven processes. This unit consists of two identical units including Haldor Topsøe's final desulfurization and steam reforming section, CO shift section, Eickmeyer & Associates' CO_2 removal section (Catacarb Process), methanation section and process condensate treating section. Hydrogen Recovery Unit which recover high purity hydrogen from refinery off gas is provided as ancillary process of this unit since this is economically justified.

4.1.3 Process design capacity

The design capacities of the process units of which the refinery will consist, will be as shown below.

			Feed Capacity (m3/hr)
CRUDE UNIT	N-RUMAILA	Unit No. 10	1,040
	MISCHRIF	Unit No. 20	1,040
N-HDS UNIT	N-RUMAILA	Unit No. 11	244
	MISCHRIF	Unit No. 21	273
CAT-REFORMER	N-RUMAILA	Unit No. 15	170
	MISCHRIF	Unit No. 25	170
K-HDS UNIT	N-RUMAILA	Unit No. 12	73
	MISCHRIF	Unit No. 22	116
GO-HDS UNIT	N-RUMAILA	Unit No. 13	102 (1)
	MISCHRIF	Unit No. 23	162
GULF HDS UNIT	N-RUMAILA	Unit No. 14	596 (2)
	MISCHRIF	Unit No. 24	497 (3)
HYDROGEN UNIT		Unit No. 02	3,800,000 Nm3/day (4)
BTX-UNIT		Unit No. 04	263 (5)
PAREX-UNIT		Unit No. 15	8.3 (6)
GAS/AMINE UNIT		Unit No. 01	(7)
SR UNIT		Unit No. 03	600 tons/day/ 2 trains

Notes:

(1) Future unit

(2) 298 m^3/hr x 2 identical unit. Each unit consists of two high pressure reactor trains in parallel.

(3) The unit consists of three high-pressure reactor trains in parallel.

(4) 1,900,000 Nm^3/day x 2 identical unit.
Hydrogen production.

(5) Depentanizer feed (Benzene production is 12 m^3/hr)

(6) Paraxylene production

(7) Just enough capacity to treat all sour gas and LPG

4.2 CRUDE UNIT

4.2.1 Process design basis

(1) Capacity and charge

All the units will be designed to be capable of processing both N-RUMAILA and MISCHRIF.

Capacities and charge stocks of the these Units are as follows:

	Unit No. 10	Unit No. 20
Crude source	N-RUMAILA	MISCHRIF
Capacity, m3/hr	1,040	1,040
(BPSD)	157,000	157,000

As to the properties of the chargestocks, please refer to the crude assay attached to Chapter 2 hereof.

(2) Product specifications

Each unit will be capable of processing the feedstocks defined above, producing the following fractions in normal operation:

Stream	TBP Cut Points on Crude, °C
Overhead distillate	To 150/175
Kerosene	150/175 - 250
Light gas oil	250 - 300
Heavy gas oil	300 - 343
Residue	343^+

When processing crude oil as specified above for normal operation the units is capable of making separations as follows between products.

(a) The ASTM distillation curves of the products corrected for loss show the following characteristics as betwen the 95% boiling point of the first product and the 5% boiling point of the second product.

Condensed overhead dis/kerosene	12°C gap min.
Kerosene/light gas oil	8°C gap min.

(b) The kerosene has a minimum closed Abel flash point of 65°C for the 175 - 250°C TBP product.

(c) The light gas oils have a minimum closed Pensky Marten flash point of 100°C.

(d) The heavy gas oils have the following characteristics:

- (i) Minimum closed Pensky Marten flash point 110°C.
- (ii) The 90% point on the ASTM distillation of the heavy gas oil shall not be more than 355°C.

(e) The residue have the minimum closed Pensky Marten flash point of 180°C.

(3) Process design notes

While the process design will conform to the TENDER DOCUMENT, Vol. II, 10.1 and 10.2.5 in principle, there will be the following deviation:

Stabilizer will not be provided and LPG plus naphtha will be directly charged to the N-HDS UNIT so that the charge is desulfurized and then stabilized.

4.2.2 Description of unit

Crude oil from the tank is boosted up by the crude charge pump. Then the oil is heat exchanged with No. 1 external reflux (top) and is split into two streams, which flow in parallel. One stream is heat exchanged with cold LGO products. The other stream is heat exchanged with kerosene products and cold HGO products. After the above exchangers, the two streams join, and is heated by heat exchanger with cold residue, extering into the desalter, where it is desalted by the electrical method.
The desalted crude oil is heat exchanged with No. 2 external reflux (kerosene) and is divided into two parallel streams.

One stream is heat exchanged with No. 3 external reflux (LGO). The other stream is heated by heat exchanger with hot LGO products and hot HGO products. Then the two streams join, and is heat exchanged with hot residue, going into the preflash drum, where it is separated into gas and liquid.

The bottoms separated in the preflash drum is introduced into the heater where it is heated to the required temperature.
After heating, the liquied joints the preflash drum vapor and is fed to the main tower.
LPG and naphtha are withdrawn from the top of the main tower. The LPG and naphtha are cooled in the air coolers and in the water cooled condenser and are introduced into the main tower overhead receiver.

Further the gas evolution from the overhead receiver is compressed, and recontacted with naphtha, entering into the N-HDS surge drum. These are then fed into the N-HDS UNIT.

Three side cuts are withdrawn from the main tower; they are kerosene, light gas oil (LGO) and heavy gas oil (HGO). The water content and the properties of kerosene are adjusted in the reboiled strippers in which residue is used as the heat source. The kerosene is heat exchanged with the crude oil and is cooled in the air coolers and in the water cooled heat exchanger, and then it is sent to storage and K-HDS UNIT.

The LGO and HGO are stripped of light components in the stripper in which steam stripping is conducted.
After heat exchange with the crude oil, the LGO is cooled in the air coolers and in the water cooled heat exchanger, entering into the salt dryer, where it is reduced to the water content of LGO. The HGO is heat exchanged with the crude oil and is cooled in the air coolers and is sent to storage and GO-HDS UNIT.

Main tower residue is used as the heat source in the kerosene stripper, and is heat exchanged with the crude oil. Then, it is cooled in the air cooler and is sent to storage and GULF HDS UNIT.

The main tower is provided with three external refluxes in the top section, at the kerosene tray and at the LGO tray. Each external reflux is heat exchanged with the crude oil, so that the properties of each product can be adjusted properly.

Besides, for stripping and removing H_2S from the foul water discharged from the overhead receiver of the PRESENT UNIT and also the various HDS UNITS, a foul water stripper of the reboiler stripping type has been provided. However, it will be located beside the SR-UNIT, instead of in the CRUDE UNIT area.

Meanwhile, please refer to the simplified process flow diagrams for the UNIT, Dwg. No. 10-G-001-D and Dwg. No. 10-G-002-D.

4.2.3 Features of unit

(1) By making the design of the N-RUMAILA train CRUDE UNIT (Unit No. 10) and that of the MISCHRIF train CRUDE UNIT (Unit No. 20) completely identical, the basic design work has been simplified and subsequently the construction cost will also be saved. This will also serve to increase operational flexibility.

(2) Deliberate considerations have been paid to the design of the heat exchangers, pumps, trays, etc. for enabling easy and stabilized 50% reduced operation.

(3) Economically optimum design arrangements between the heat exchanger network for crude oil preheating and heat exchanger size has been contrived.

(4) The desalter has been designed to be of the two stage type as required by the GULF HDS UNIT that is located subsequent to the desalter.

(5) Preflash drum has been provided at the outlet of the final crude preheating exchanger. Due to this, vaporization of crude oil at the heater inlet can be prevented and this has enabled lower design pressures of the preheat exchangers and desalter.

(6) While the trays have been designed generally to be of the valve type, those from the flash zone to the HGO draw-off tray have been designed as bubble cap trays so that coloring of HGO and mixture of metal content are prevented.

(7) It has been so designed that the full amount of gas that evolves from the overhead receiver, after having been pressurized by means of a compressor, will be recovered through gas/naphtha recontacting by being dissolved into naphtha.

(8) A stabilizer will not be provided in the CRUDE UNIT and the naphtha with gas dissolved in it as per above will be directly fed to the N-HDS UNIT so as to be desulfurized and when stabilized. This is for avoiding uneconomical arrangement whereby naphtha stabilization is doubled and also making LPG mercaptan-sulfur-free.

(9) As a part of the kerosene will be blended with product without desulfurization, for the purpose of conforming the water content to the specification, the kerosene side cut stripper has been designed to be of the reboiled stripper type. Heat source of the reboiler will be reduced crude.

(10) LGO and HGO will be stripped with steam that has been superheated in the heater convection section so that their flash points, etc. are adequately adjusted. As LGO may be blendied with products without desulfurization, its water content will be reduced through a salt dryer after its temperature has been cooled.

(11) Reduced crude, also like LGO and HGO, will be stripped with superheated steam.

(12) Two foul water strippers with equal capacity will be provided so that fluctuations in feed rate can be coped with and flexibility for reduced operation is available.

4.2.4 Estimated yield and properties

(1) N-RUMAILA crude

The crude assay attached is a revised one prepared by us through recalculation. Yields and their properties with the PRESENT CRUDE UNIT indicated hereunder are based on the revised assay.

	Crude	LPG+ Naphtha	Kerosene	LGO	HGO	Residue
Vol% on crude	100	22.0	11.0	7.1	5.6	54.3
m^3/hr	1,040	229	114	74	58	565
Sp.Gr. (15/15°C)	0.868	0.702	0.794	0.837	0.854	0.957
Sulfur content (wt%)	2.19	0.005	0.15	0.96	1.60	3.38
Flash point (°C)	-	-	65	104	135	210
ASTM distil (°C)						
IBP	-	-	149	186	220	-
5 vol%	-	-	192	256	294	-
10	-	-	194	258	299	-
30	-	-	203	267	312	-
50	-	-	213	274	321	-
70	-	128	223	282	332	-
90	-	152	237	296	351	-
95	-	163	244	301	356	-
EP	-	181	272	344	406	-
ASTM 5 to 95% gap (°C)		+29	+12	-7		

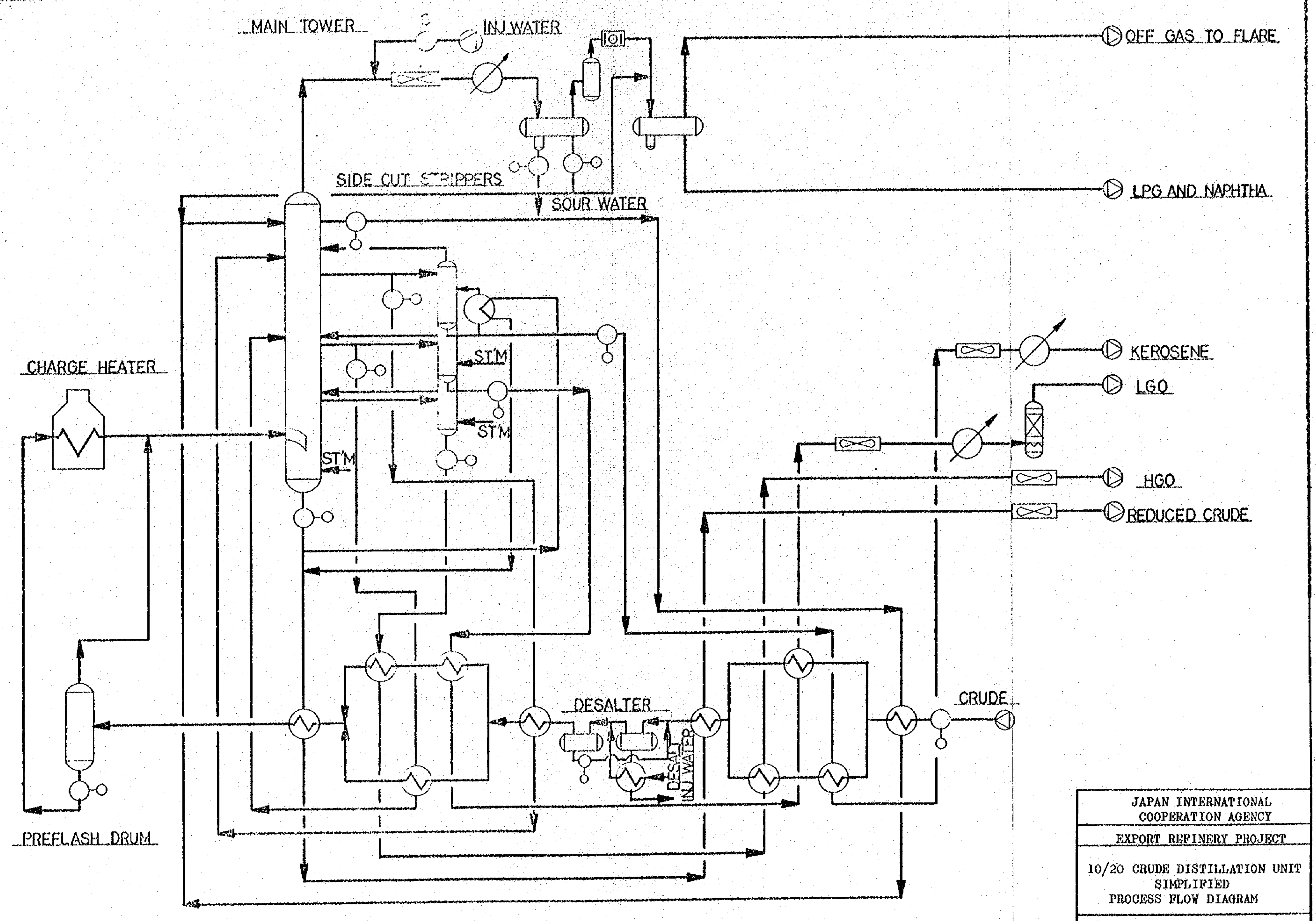
(2) MISCHRIF crude

In view of the fact that the crude assay attached to the TENDER DOCUMENTS is of such simple form that lacks graph of Specific Gravity vs TBP Distillation, yields and their properties based on the results of calculations conducted by us are indicated hereunder for your information.

	Crude	LPG+ Naphtha	Kerosene	LGO	HGO	Residue
Vol% on crude	100	24.5	10.2	7.6	7.2	50.5
m^3/hr	1,040	255	106	79	75	525
Sp.Gr. (15/15°C)	0.8795	0.701	0.803	0.848	0.878	0.992
Sulfur content (wt%)	3.88	0.01	0.59	1.95	2.94	6.07
Flash point (°C)	-	-	65	106	134	224
ASTM distil (°C)						
IBP	-	-	159	188	220	-
5 vol%	-	-	197	256	288	-
10	-	-	200	257	294	-
30	-	-	210	265	309	-
50	-	-	221	270	317	-
70	-	130	229	277	324	-
90	-	156	241	292	346	-
95	-	168	248	298	352	-
EP	-	183	273	343	397	-
ASTM 5 to 95% gap (°C)		+29	+8	-10		

4.2.5 Battery limit conditions

Fluid	Temperature (°C)	Pressure (kg/cm2G)	Note
Crude	40	1.18	
Naphtha	50	43.5	Direct charge
Kerosene	88	5.0	Direct charge
	40	5.0	Rundown
LGO	93	5.0	Direct charge
	38	5.0	Rundown
HGO	113	5.0	Direct charge
	65	5.0	Rundown
Reduced crude	144	5.0	Direct charge
	80	5.0	Rundown
Desalter effluent water	60	3.0	



JAPAN INTERNATIONAL
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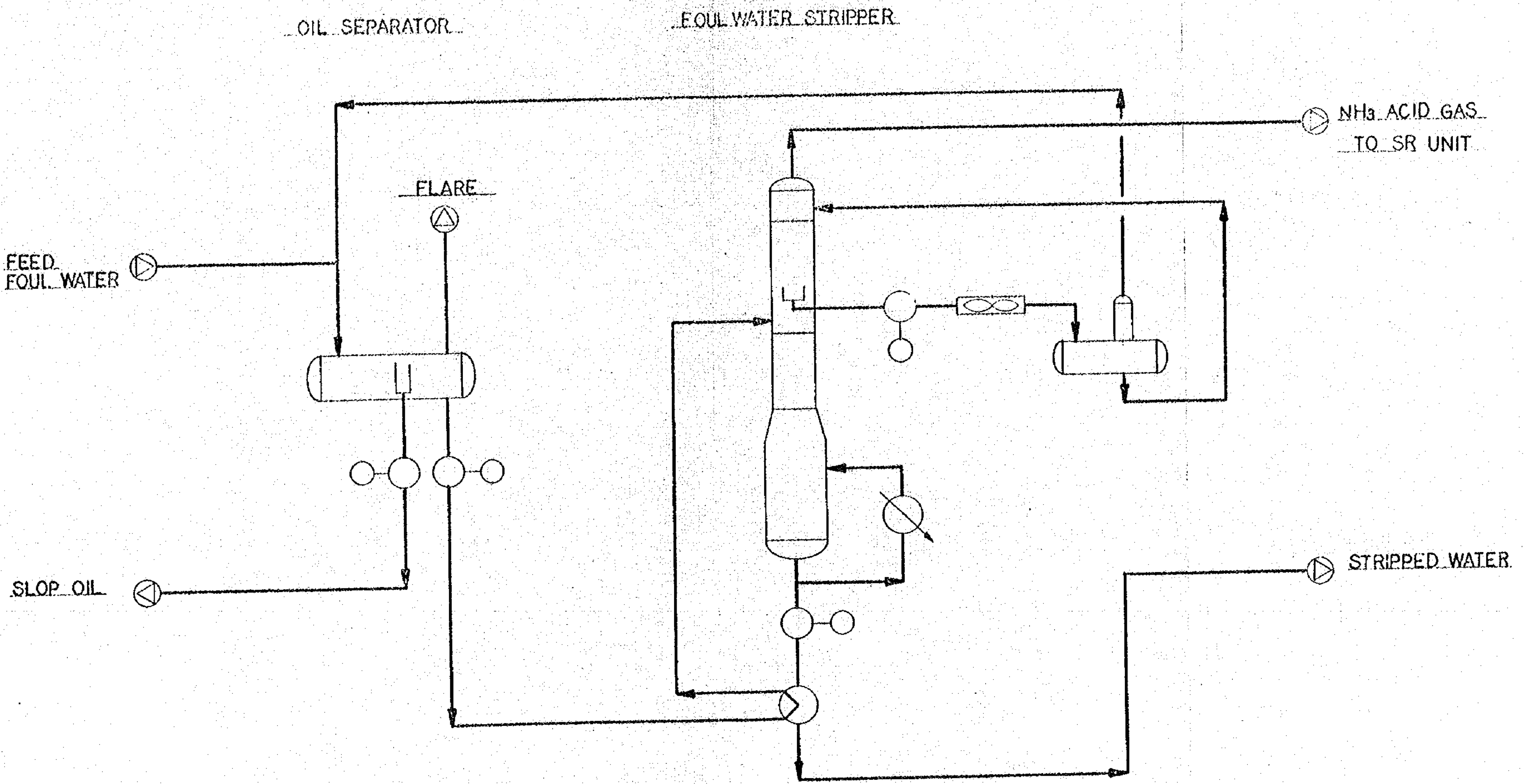
EXPORT REFINERY PROJECT

10/20 CRUDE DISTILLATION UNIT
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 10-G-001



JAPAN INTERNATIONAL
COOPERATION AGENCY

EXPORT REFINERY PROJECT

10/20
FOUL WATER TREATING SECTION
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 10-G-002

4.3 N-HDS UNIT

4.3.1 Process design basis

(1) Capacity and charge

Capacity and chargestock of these unit are as follows:

	Unit No. 11	Unit No. 21
Crude source	N-RUMAILA	MISCHRIF
TBP cut point, °C	To 150/175	To 150/175
Capacity, m^3/hr	244	273
(BPSD)	(36,800)	(41,200)

Design is based on 175°C TBP cut naphtha.
Naphtha withdrawn as overhead distillate from the CRUDE UNITS will be charged to the N-HDS UNITS: naphtha produced from N-RUMAILA will be charged to Unit No. 11 and naphtha produced from MISCHRIF will be changed to Unit No. 21.

(2) Product specification

To hydrotreat the feedstock using H_2 rich gas from either of the CAT-REFORMERS.
To stabilize the treated effluent and to fractionate this given the stabilized gasoline and naphtha products as follows:

	TBP Cut Points (°C)
Off-gas	-
C_4 and lighter	-
LSR gasoline	C_5 - 65
Naphtha	65 - 150/175

All products heavier than butane are doctor negative and free of H_2S and elemental sulfur; in addition, the products meet the following:

(a) C_4 and lighter

Pentanes content, vol% max. : 1.0

(b) LSR gasoline

(i) C_4 and lighter content, vol% max.: 1.0

(ii) Copper strip corrosion, max. : No. 1

(iii) Total sulphur content, ppm : 0.5

(iv) The 95/5 vol% gap in the ASTM distillations with the naphtha product, °C min. : 10

(c) Naphtha

(i) Total sulphur content, wt ppm : 0.5

(ii) Copper strip corrosion, max. : No. 1

(3) Process design notes

TENDER DOCUMENT Vol. II 10.1 and 10.3.5 will be complied with except for the two deviations explained below.

(a) LPG plus naphtha will be charged to this unit without being stabilized through the CRUDE UNIT. During normal operation, therefore, consideration is given to direct charging from the CRUDE UNIT only.

(b) Hydrotreated naphtha of 150 - 175°C TBP cut will not be produced which reduces the number of splitters to one. As this fraction is planned to be blended with aviation turbine fuel, a slight amount of 65 - 175°C TBP cut will be used instead.

4.3.2 Description of UNIT

The process flow is illustrated in the attached simplified flow diagram; Dwg. No. 11-G001-D.

The charge stock, together with the make-up hydrogen and recycle gas, are heated to reaction temperature in exchangers and a fired heater. In the reactor the contained sulphur, nitrogen, halogen and oxygen impurities, together with olefinic hydrocarbons, are converted over hydrobon catalysts. All the above reactions are exothermic, although the exothermic heat is relatively small when minimal quantities of these impurities are present.

The reaction product is cooled through exchangers and a cooler on its way to the product separator. Net separator gas is recycled and combined with hydrogen rich make-up gas while the separator liquid, after heat exchange, is stabilized in the stabilizer column to remove H_2S and sour light ends.

H_2S rich sour gas is introduced into the GAS/AMINE UNIT low pressure gas scrubber, while the sour light ends, withdrawn as stabilizer overhead liquid product, are sent to the GAS/AMINE UNIT LPG extractor.
The stabilizer column bottoms product, after heat exchange, is split in the splitter column to separate LSR gasoline and naphtha.

The LSR gasoline, withdrawn as overhead liquid product, is sent to storage. The naphtha is withdrawn from bottoms of the splitter, heat exchanged and/or cooled.
Then the naphtha is fed to the CAT-REFORMER or the storage tank in the off-site area.

4.3.3 Features of process

(1) Catalysts can be regenerated using a steam-air procedure.

(2) Most of the separator gas is recycled to minimize the quantity of the make-up hydrogen.

4.3.4 Estimated yield and properties

(1) N-RUMAILA naphtha

Chargestock: 36,800 BSPD gas 175°C TBP naphtha

Properties : 70.0 °API, 50 wt ppm sulfur, 1.0 wt ppm nitrogen

	Wt%	SG	Vol% (SCFB)	BPSD
Feedstock:				
Naphtha feedstock	100.00	0.702	100.00	36,800
Hydrogen (Chemical)	0.05		(24)	
Net products:				
Hydrogen sulfide	0.01			
Methane and ethane	0.01			
Propane produced	0.01			
Butanes produced	0.02	0.579	0.02	7
Naphtha product (a)	100.00	0.702	100.04	36,815
Total:	100.05	-	100.06	36.822

Product: 70.1 °API, 0.5 wt ppm sulfur maximum

Note: (a) Includes all C_2, C_3 and C_4 in the feedstock.

(2) MISCHRIF naphtha

Chargestock: 41,200 BPSD gas 175°C TBP

Properties : 70.3 °API, 0.01 wt% sulfur, 1.0 wt ppm nitrogen

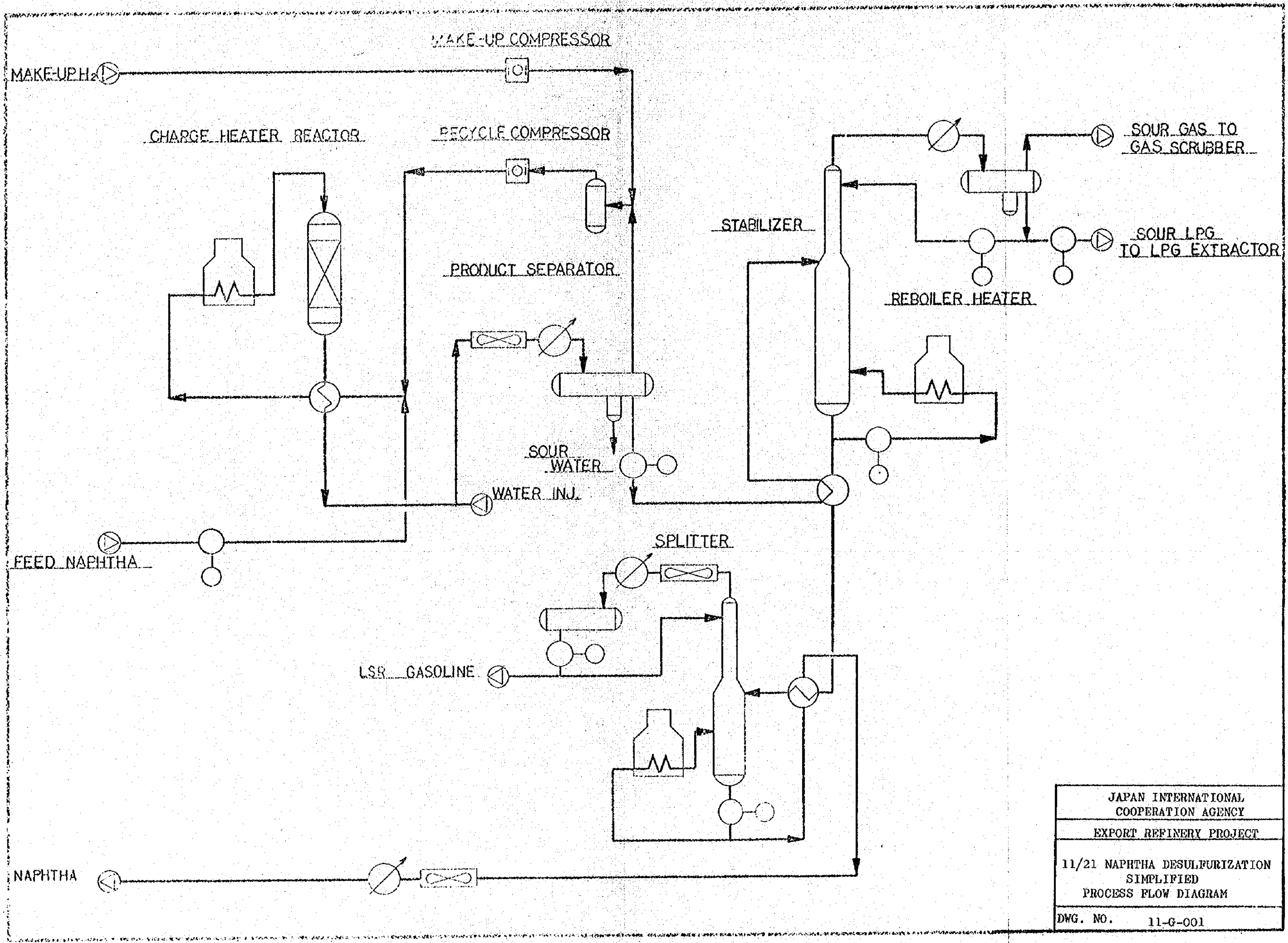
	Wt%	SG	Vol% (SCFB)	BPSD
Feedstock:				
Naphtha feedstock	100.00	0.701	100.00	41,200
Hydrogen (Chemical)	0.07		(30)	
Net products:				
Hydrogen sulfide	0.11			
Methane and ethane	0.02			
Propane produced	0.02			
Butanes produced	0.02	0.579	0.03	12
Naphtha product (a)	99.90	0.700	100.04	41,217
Total:	100.07	0	100.07	41,229

Product: 70.6 °API, 0.5 wt ppm sulfur maximum

Note: (a) Includes all C_2, C_3 and C_4 in the feedstock

4.3.5 Battery limit conditions

Fluid	Temperature (°C)	Pressure (kg/cm^2G)	Note
Full range naphtha LPG	50	43.5	
Distillate light Naphtha	40	5.0	
Distillate heavy naphtha	40	5.0	
Distillate heavy naphtha	90	6.5	To CAT-REFORMER
Slop header	40	5.0	
Start up bypass	40	5.0	From CAT-REFORMER
DHN (Dilution of Chemical)	40	3.0	To CAT-REFORMER
Waste water	40	3.0	
Treated water	40	3.0	
Electrolytic hydrogen of recycles gas	40	12.2	CAT-REFORMER or HYDROGEN UNIT
LPG (Sour)	40	16.0	
Comp. suct. drum drain	40	7.75	To CAT-REFORMER
H_2 rich gas	40	12.2	From CAT-REFORMER
H_2 gas	40	12.2	To CAT-REFORMER
Stab. OVHD gas	40	6.3	

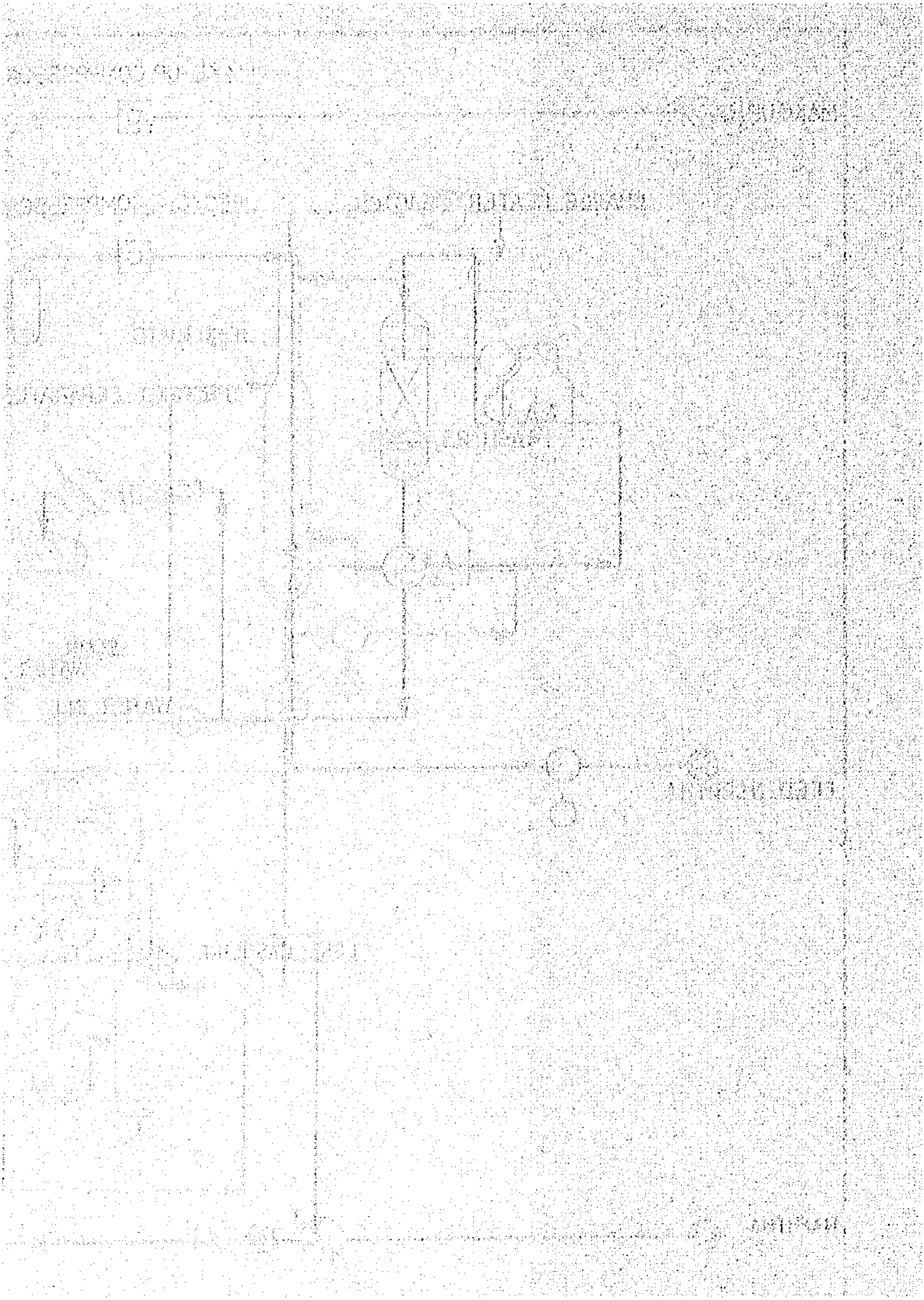


JAPAN INTERNATIONAL
COOPERATION AGENCY

EXPORT REFINERY PROJECT

11/21 NAPHTHA DESULFURIZATION
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 11-G-001



4.4 CAT-REFORMER UNIT

4.4.1 Process design basis

(1) Capacity and charge

The unit of the N-RUMAILA train is designated Unit No. 15 and that of the MISCHRIF train, Unit No. 25. Both Units will have the same capacity and be designed so that hydrotreated naphtha obtained from either crude can be handled.
The capacity and chargestock concerned are given below.

	Unit No. 15 or 25	
TBP cut point °C	65 - 150	65 - 175
Capacity, m^3/hr	129	170
(BPSD)	(19,500)	(25,700)

Design case will be 65 - 175°C TBP cut.
The data in the TENDER DOCUMENTS for N-RUMAILA naphtha with an endpoint of 175°C and for MISCHRIF naphtha in general are poor, and would not be suitable as the basis of a final design. The same design would be appropriate for both the N-RUMAILA and MISCHRIF Platforming Units since the naphthas are so very similar in terms of their specific gravity and the ratio of paraffins, naphthenes and aromatics.
For this the simplification has therefore been made that the N-RUMAILA and MISCHRIF naphthas can both be characterised by the same figures for the 65 - 175°C TBP and 65 - 150°C TBP materials.

(2) Products specification

The unit is designed to produced a maximum yield of stabilized reformate and hydrogen at severities up to 98 (CFR research method, F1, clear) octane number from either of the above feedstocks.

(a) Reformate

(i) Research octane number clear : 98

(ii) C_4 and lighter content, vol% max. : 2.0

(iii) Copper corrosion test, max. : No. 1

(b) LPG

(i) C_5 content, vol% max. : 1.0

(ii) Copper corrosion test, max. : No. 1

(3) Process design notes

These will conform to TENDER DOCUMENT Vol. II, 10.1 and 10.4.5.

4.4.2 Description of UNIT

(1) Reactor section

A simplified flow diagram of this unit is contained in the attached drawing; Dwg. No. 15-G-001-D.

The basic flow pattern through the UOP Continuous Platformer is essentially the same as with conventional fixed bed Units. Feed to the platformer is combined with recycle hydrogen, raised to the reaction temperature by heat exchange and a fired heater and charged to the reactor section. Temperature is maintained across the reaction section by interheaters. Effluent from the reactors is cooled by heat exchange and subsequent air and water cooling, and is charged to the separator section. Separator gas is recycled to the reactors with the net hydrogen make sent to hydrogen users. Separator liquid is directed to product stabilization facilities.

The catalyst can move through the reactors to the regeneration section where a small quantity is continually being regenerated.

For this operation the platforming reactors are stacked one on top of another and this results in a very compact unit with a minimum of piping. The reactor internals, are, in principle, similar to a conventional platforming reactor. The catalyst moves in essentially plug flow down through each reactor and through reactor connection lines to the catalyst regeneration system.

(2) Catalyst regeneration section

The catalyst flows by gravity from the last reactor into a catalyst collector vessel which automatically maintains uniform catalyst withdrawal. After depressuring and purging, the catalyst gravity flows to a simple lift engager that pneumatically lifts the catalyst to a disengaging hopper.

The catalyst then gravity flows to the regeneration tower where the carbon is burned off. The net regeneration gas is vented. The catalyst is lifted to the top of the reactor stack and is reduced in the top section before reentering the processing zone of the reactors.

The flow of catalyst is carefully controlled through the lock hoppers and lift engagers to prevent any catalyst attrition.

The regeneration and reactor sections of the unit are easily isolated.

4.4.3 Features of process

(1) The continuous regeneration concept permits plant operation at very low operating pressures leading to maximization of hydrogen and reformate yields.

(2) Hydrogen production is continuous and consistent in composition.

(3) The plant design is simpler and thus less costly than the cyclic system which involves comples valving systems and an additional "Swing" reactor.

4.4.4 Estimated yield and properties

(1) 65 - 175°C TBP naphtha

Chargestock: 25,700 BPSD 65 - 175°C TBP naphtha from N-RUMAILA or MISCHRIF

Properties : 60.0 °API, 0.5 wt ppm sulphur max., 68/21/11 vol% ratio of paraffins/naphthenes/aromatics

	Wt%	SG	Vol% (SCFB)	BPSD
Feedstock:				
Naphtha	100.00	0.739	100.00	25,700
Product:				
Hydrogen	2.60		(1,266)	
Methane	1.08			
Ethane	2.66			
Propane	3.70			
I-butane	1.81	0.563	2.38	612
N-butane	2.72	0.584	3.44	884
I-pentane	2.94	0.625	3.48	894
N-pentane	1.96	0.631	2.29	589
Hexanes and all heavier material	80.53	0.811	73.41	18,866
Total:	100.00	-	85.00	21,845

De-butaniser bottoms product: 19,840 BPSD

Properties: 45.5 °API, 3.77 psi RVP, 98.0 RON clear

(2) 65 - 150°C TBP naphtha

Chargestock: 19,500 BPSD 65 - 150°C TBP naphtha

Properties : 63.0 °API, 0.5 wt ppm sulphur max., 74/18/8 vol% ratio of paraffins/naphthenes/aromatics

	Wt%	SG	Vol% (SCFB)	BPSD
Feedstock:				
Naphtha	100.00	0.727	100.00	19,500
Product:				
Hydrogen	2.55		(1,223)	
Methane	1.35			
Ethane	4.27			
Propane	5.93			
I-butane	2.89	0.563	3.73	727
N-butane	4.33	0.584	5.39	1,051
I-pentane	4.46	0.625	5.19	1,012
N-pentane	2.97	0.631	3.42	667
Hexanes and all heavier material	71.25	0.809	64.01	12,482
Total:	100.00	-	81.74	15,939

De-butaniser bottoms product: 13,630 BPSD

Properties: 47.1 °API, 4.67 psi RVP, 98.0 RON clear

(3) Preliminary break-up of hydrocarbon types

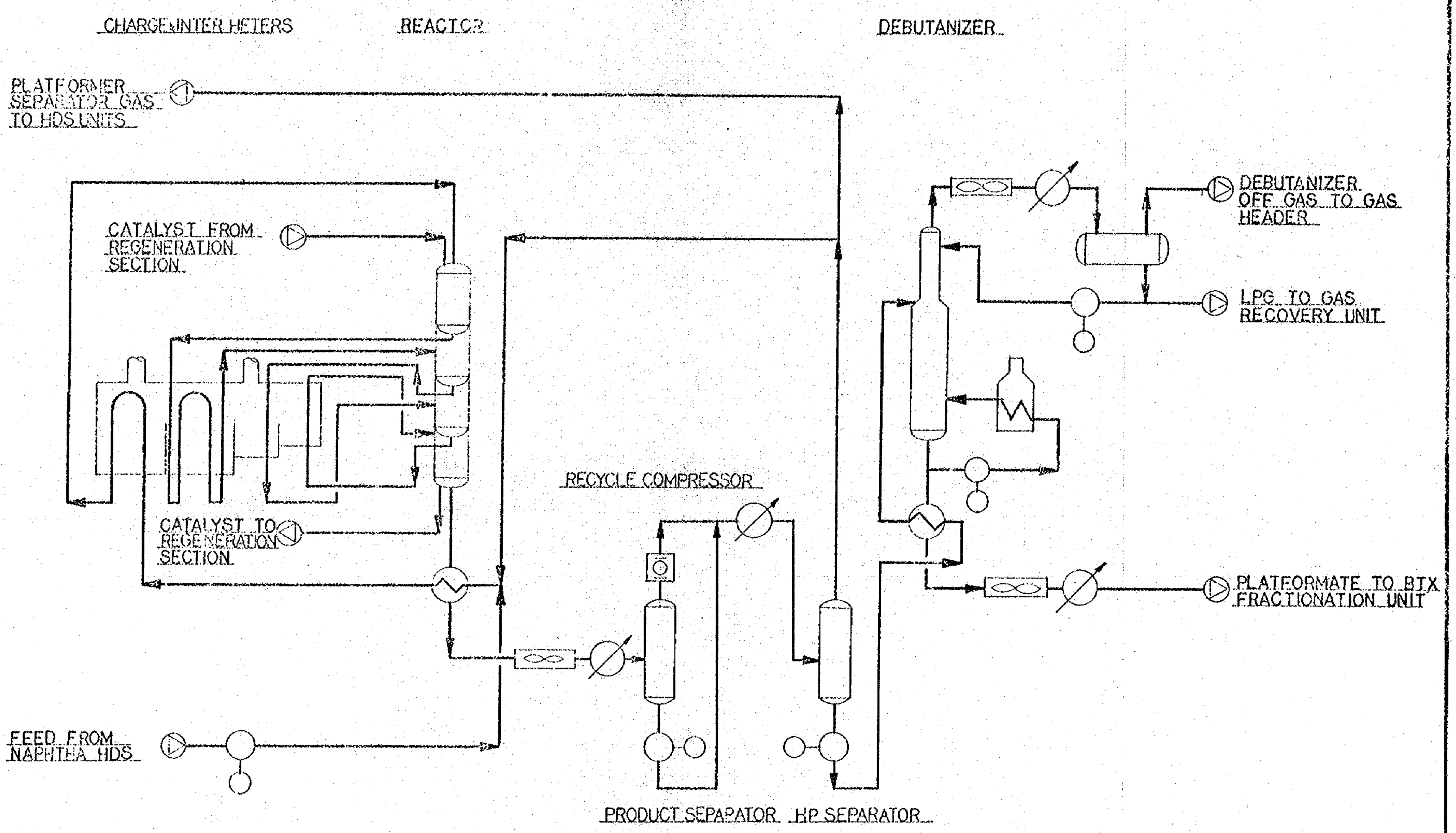
Naphtha chargestock:

	LV%	BPSD	LV%	BPSD
Chargestock quantity:	100.0	25,700	100.0	19,500
De-butaniser bottoms:	77.2	19,840	69.9	13,630
Component break-up :				
Normal butane	0.6	155	0.6	115
Iso pentane	2.7	705	4.1	790
Normal pentane	1.9	490	2.8	550
Iso hexanes	4.7	1,215	8.8	1,720
Normal hexane	2.6	655	4.8	930
Methyl cyclopentane	0.3	65	0.1	20
Benzene	3.6	920	4.8	945
Iso heptanes	6.7	1,710	2.7	520
C_7 napthenes	0.1	30	0.4	90
Normal heptane	2.8	735	1.1	220
Toluene	9.6	2,480	13.8	2,685
Iso octanes	3.7	945	0.2	40
C_8 naphthenes	0.1	25	0.2	35
Normal octane	1.2	310	0.1	15
Ethyl benzene	3.0	780	3.2	635
Paraxylene	2.8	720	3.2	625
Metaxylene	5.8	1,490	6.9	1,345
Orthoxylene	3.8	980	4.6	890
C_9 saturates	1.4	360	0.2	45
Isopropyl benzene	0.2	50	0.2	30

	LV%	BPSD	LV%	BPSD
N-propyl benzene	0.6	150	0.4	80
Methyl ethyl benzenes	3.7	940	2.7	525
Trimethyl benzenes	5.4	1,390	4.0	780
C_{10} aromatics	9.9	2,540	-	-

4.4.5 Battery limit conditions

Fluid	Temperature (°C)	Pressure (kg/cm^2G)	Note
Distillated heavy naphtha	40	6.5	
Distillated heavy naphtha	90	8.0	From N-HDS UNIT
Reformate	40	5.0	
Start-up bypass	40	5.0	
LPG (Sweet)	40	16.0	
H_2 rich gas	40	12.2	
H_2 rich gas	40	14.1	
H_2 gas	40	12.2	From N-HDS UNIT
Deb. OVHD gas	40	6.0	



JAPAN INTERNATIONAL
COOPERATION AGENCY

EXPORT REFINERY PROJECT

15/25 CATALYTIC REFORMER
(UOP C.C.R. PLATFORMER)
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 15-G-001

4.5 BTX UNIT

4.5.1 Process design basis

(1) Charge and capacity

This UNIT will be capable of processing all reformate to be received from both Units Nos. 15 and 25.

	Capacity	
	m^3/hr	(BPSD)
Feed rate to depentanizer (a)	263	(39,680)
Feed rate ot extraction section	54	(8,162)
Production rate of benzene	12.1	(1,827)

Note: (a) The feed streams are initially processed in this column.

(2) Product specification

The quality of the benzene product is as described below.

Benzene content : 99.8 vol% min.

Toluene : 0.1 vol% max.

Non-aromatics : 0.1 vol% max.

Sulphur : 1 wt ppm max.

Thiophene : 1 wt ppm max.

H_2S/SO_2 : NIL

Acid wash colour : 1 max.

Residue on evaporation: 5 mg/100 ml max.

Appearance : Clear; no suspended matter

Boiling range : 1°C max., including 80.1°C

Freeze point : 5.45°C min.

Acidity : No free acid

Benzene recovery : 99.0 vol%

(3) Process design Notes

Vol. II. 10.1, 10.5.3 and 10.5.4 of the TENDER DOCUMENT will be complied with.

4.5.2 Description of UNIT

(1) Fractionation section

A simplified flow diagram of this section is given in the attached drawing; Dwg. No. 04-G001-D.

The operation of the de-pentaniser has been designed such that not only the pentanes but also half the isohexanes pass overhead and thus go directly to the gasoline pool whilst 99.6% or more of the benzene goes on to the splitter.
For the splitter the operation is such that approximately 99.6% of the benzene but only about 1.0% of the toluene in the feed to the column go into the overhead product.
This product is then the feed to the SULFOLANE PROCESS UNIT.
The bottoms product of the splitter goes on to the de-heptaniser which has been designed to obtain the majority of the toluene as the overhead product ready for blending directly to the gasoline pool. The bottoms product is to contain only 1% of the toluene but more than 99.8% of the paraxylene in the total platformate. The bottoms product is then clay treated.

After hot clay treatment, the heavy portion of the combined platformate is to be fed to a xylenes rerun column which not only separates out the heavy material formed during the clay treatment but also prepares a UOP PAREX PROCESS UNIT feedstock containing only 500 molar ppm of C_9 aromatics.
The advantage of this is that the extracted paraxylene from the UOP PAREX UNIT does not need to be rurun to meet the specification for C_9 aromatics. The recovery of paraxylene overhead in this column has been set such as to meet the requirements given in the TENDER DOCUMENTS which ask for only 8.2 m^3/hr of finished paraxylene product. For the high endpoint platformate 8.6 m^3/hr could in fact be produced, but for the low endpoint only 7.5 m^3/hr would be possible at the design throughputs.

(2) BTX extraction section

A simplified flow diagram of the SULFOLANE PROCESS UNIT is given in the attached drawing; Dwg. No. 04-G002-D.

The hydrocarbon feed is pumped to extraction column and is counter-currently contacted with sulfolane solvent. As the feed flows through the extractor, aromatics are selectively dissolved in the solvent, and raffinate of low aromatics content is withdrawn from the top of the extractor.

The rich solvent from the extractor is sent to a stripping section which consists of a stripper and extract recovery column. The stripper column operates at a pressure which is high enough to prevent flashing of the rich solvent.
The extract, from the bottom of the stripper, essentially contains only aromatics and solvent. This stream is sent to the recovery column where the aromatics are removed from the solvent by steam stripping. The extract product is then directed to the clay treating and fractionation facilities.
The water from the recovery column overhead receiver is returned to the raffinate water wash column. Most of the lean solvent from the recovery column is returned to the extractor. The regenerated solvent is put back into circulation via the recovery column. Degradation products resulting from any decomposition of the solvent and other undesirable material are removed in the solvent regenerator.
Lean solvent from the recovery column bottoms is used to reboil the water stripper. Lean solvent is also heat exchanged with rich solvent before being returned to the top of the extractor.

Raffinate, the non-aromatic hydrocarbon phase, leave the top of the extractor. Recovery of the small amounts of sulfolane solvent contained in the raffinate stream takes place in the water wash column. The sulfolane-free raffinate is then sent to battery limits.

The wash water is charged to the water stripping column where it is stripped free of traces of dissolved paraffins before being returned to the extract recovery column.

The extracted material then needs to be hot clay treated and separated from the heavies produced and the toluene in a rerun column from which dry specification benzene product can be taken as a sidecut from high up the column.

4.5.3 Features of process

The Shell Sulfolance Process is a commercially proven solvent extraction process for recovery of high purity aromatics, particularly benzene, toluene and xylenes from hydrocarbon mixtures. Liquid-liquid extraction and extractive stripping techniques are combined with the use of sulfolane solvent to obtain high purities and recoveries of aromatic components at low investment and operating costs. Reduced investment and operation costs are direct results of the low solvent ratios that are possible with the high solvency and selectivity that are characteristic of sulfolane.

4.5.4 Estimated yields and properties

(1) Fractionation section of platformate

(a) Platformate from 65 - 175°C TBP naphtha case

In	BPSD
Feed to the depentanizer (Platformate)	39,680
Out	
Depentanizer overheads to storage (Gasoline blending)	3,934
Splitter overheads to sulfolane unit	8,162
Deheptanizer overheads to storage (Gasoline blending)	8,527
Xylene rerun overheads to PAREX UNIT	7,864
Xylene rerun bottoms to storage (Gasoline blending)	11,193

(b) Platformate from 65 - 150°C TBP cut naphtha case

In	BPSD
Feed to the depentanizer (Platformate)	27,260
Out	
Depentanizer overheads to storage (Gasoline blending)	4,694
Splitter overheads to sulfolane unit	7,120
Deheptanizer overheads to storage (Gasoline blending)	5,486
Xylene rerun overheads to PAREX UNIT	6,476
Xylene rerun bottoms to storage (Gasoline blending)	3,484

(2) BTX extraction section

The following table shows the yield of the sulfolane process when using splitter (dehexanizer) column overhead product as chargestock.

Chargestock: 8,162 BPSD splitter column overhead

Operation : 99.7% recovery of 99.8% purity benzene

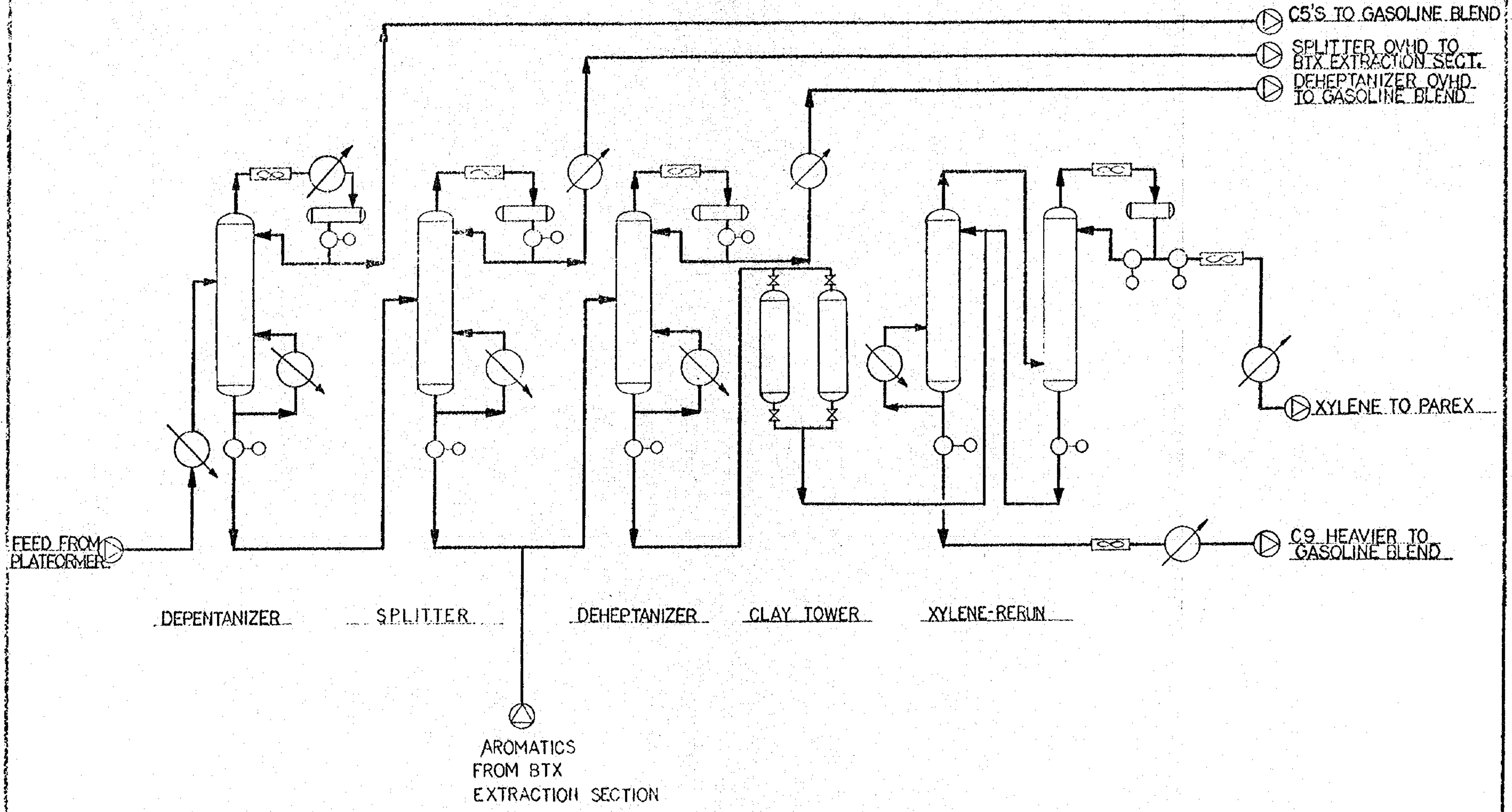
Component: BPSD	Feed	Extract	Raffinate
C_5 Paraffins	2	-	2
C_6 Paraffins	2,525	-	2,525
C_6 Naphthenes	130	-	130
Benzene	1,829	1,824	5
C_7 Paraffins	3,595	1	3,594
C_7 Naphthenes	50	-	50
Toluene	31	31	-
Total:	8,162	1,856	6,306

N.B. Benzene product is obtained by clay treating and re-running the extracted material.

The toluene rich rerun column bottoms is to be returned to the de-heptanizer feed.

4.5.8 Battery limit conditions

Fluid	Temperature (°C)	Pressure (kg/cm^{2}G)	Note
Reformate	38	5.0	From at reformer or storage tank
Heptane	38	5.0	
Mixed xylene	60	5.0	
Raffinate	38	5.0	
Benzene	43	5.0	
Off spec.' reformate	40	5.0	
Off spec.' reformate	38	5.0	From reformate storage tank
Heavy aroma	43	5.0	To heavy aroma storage including from PAREX UNIT
Heavy aroma	43	5.0	To LSRG storage tank including from PAREX UNIT

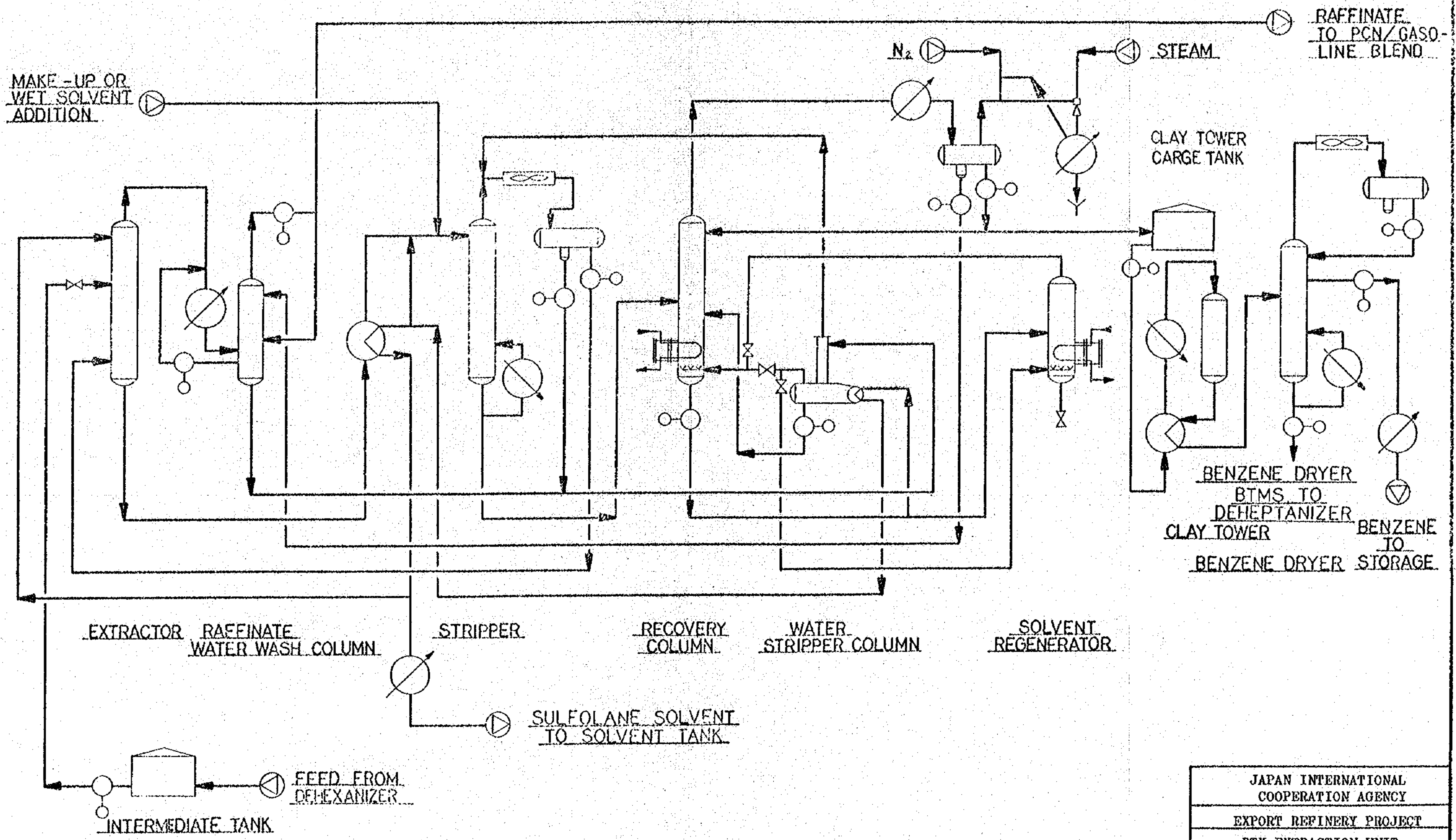


JAPAN INTERNATIONAL
COOPERATION AGENCY

EXPORT REFINERY PROJECT

BTX EXTRACTION UNIT
(FRACTIONATOR SECTION)
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 04-G-001



JAPAN INTERNATIONAL COOPERATION AGENCY

EXPORT REFINERY PROJECT

BTX EXTRACTION UNIT
(EXTRACTION SECTION)
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 04-G-002

4.6 PAREX UNIT

4.6.1 Process design basis

(1) Capacity and charge

Receiving all the overhead liquid product from the Xylene Rerun Column, this UNIT will be capable of recovering paraxylene at the specified recovery rate (See note).

	Capacity m^3/hr	(BPSD)
Feed rate to this UNIT	52.1	(7,864)
Production rate of paraxylene	8.2	(1,244)

Note: Recovery rate of 95% specified in the TENDER DOCUMENT is changed to 92% per pass for economical reasons.

(2) Product specification

The quality of the paraxylene product is described below.

Paraxylene content	: 99.0 vol% min.
Non-aromatics	: 0.2 vol% max.
Other xylenes	: 0.8 vol% max.
Boiling range	: 1°C max., including 138.4°C
Acid wash colour	: 2 max.
H_2S/SO_2	: Nil
SG 15/4	: 0.865 - 0.866
Appearance	: Clear; no suspended matter
Paraxylene recovery	: 92 vol%

(3) Process design note

TENDOER DOCUMENTS Vol. II. 10.1 and 10.6.2 will be complied with except as mentioned in Note of Para. 1) above.

4.6.2 Description of process

A simplified flow diagram of this unit is given in the attached drawing; Dwg. No. 05-G001-D.

The molecular sieve adsorbent is arranged as a succession of fixed beds with provision for the distribution and collection of inlet and outlet streams of hydrocarbon between each of the several beds of adsorbent.

4.6.3 Features of process

(1) The parex process is designed to recover more than 90% of the paraxylene in the feed in a simple once-through operation, and requires no system of refigeration or of solids handling.

(2) When maximum paraxylene production from mixed xylenes is desired and the process is therefore operated together with an isomerization unit such as a UOP ISOMAR PROCESS UNIT, the investment and operating costs of the isomerization and fractionation steps are very much lower due to the high paraxylene recovery per pass and the reduced recycle rate.

(3) Mild operating conditions and the absence of corrosive catalyst or chemicals make it possible to maximise the use of carbon steel as the construction material and this reduces the investment required.

(4) Due to the simplicity of the operation and the essentially inert environment in which the process is carried out, maintenance costs are low and minimum operator attendance is required.

(5) For the typical refinery the UOP parex process provides an economical system for the recovery of para-xylene and uses conventional equipment that is completely familiar to refinery operating personnel.

4.6.4 Estimated yield and properties

The following table shows the yield when platformate of 65 ~ 175°C TBP cut naphtha is used as the chargestock.

Chargestock: 7,864 BPSD xylenes rerun column overheads

Operation : 92% recovery per pass of 99.4% purity paraxylene

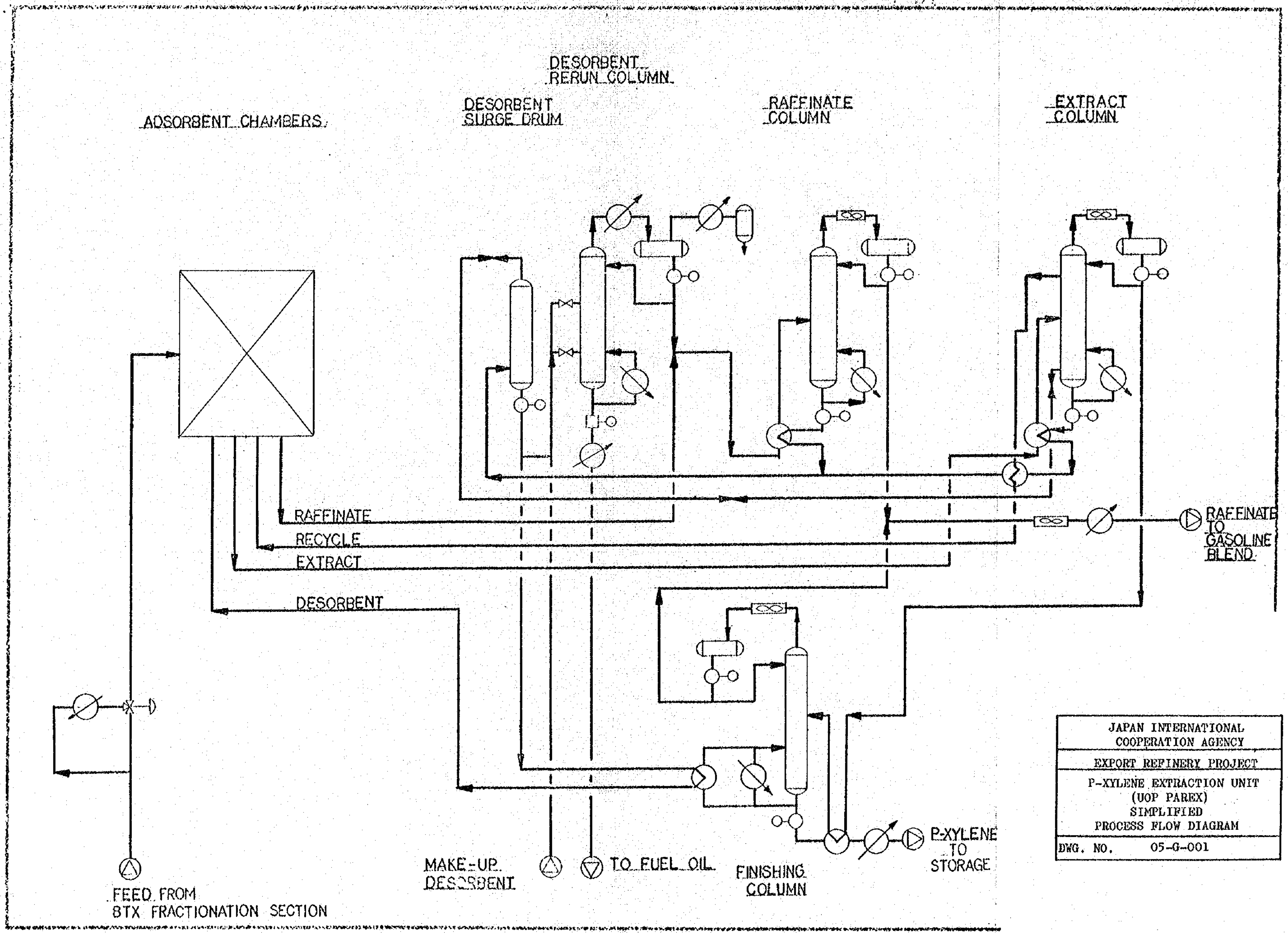
Component: BPSD	Feed	Extract	Raffinate
Toluene	50	12	38
Saturated C_8 and C_9	910	-	910
Paraxylene	1,345	1,237	108
Other C_8 aromatics	5,555	8	5,547
Isopropyl benzene	4	-	4
Total:	7,864	1,257	6,607

N.B. This material balance is an overall balance and not a description of the streams leaving the sieve chambers.

The extract stream in fact contains additional material which is separated from the desorbent and used as extract reflux.
Paraxylene product is obtained by stripping the toluene out of the extract.

4.6.5 Battery limit conditions

Fluid	Temperature (°C)	Pressure (kg/cm^2G)	Note
Mixed xylene	43	20	
Fuel oil	43	5.0	
Paraxylene	60	5.0	



JAPAN INTERNATIONAL
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EXPORT REFINERY PROJECT

P-XYLENE EXTRACTION UNIT
(UOP PAREX)
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 05-G-001

4.7 K-HDS UNIT

4.7.1 Process design basis

(1) Capacity and charge

The kerosene streams from the CRUDE UNITS will be charged to the respective K-HDS UNITS: kerosene produced from N-RUMAILA will be charged to Unit No. 12 and kerosene produced from MISCHRIF to Unit No. 22.

Capacity and chargestock of these UNITS are as follows:

	Unit No. 12	Unit No. 22
Crude source	N-RUMAILA	MISCHRIF
TBP cut point, °C	175 - 250	175 - 250
Capacity, m^3/hr	73	116
(BPSD)	(11,000)	(17,500)

(2) Product specification

To treat the feedstocks utilizing the hydrogen-rich gas from the CAT-REFORMERS as make-up and strip the reactor effluents into off-gases, naphtha and lighter streams and the treated kerosene products.

The stripped product is free of H_2S and elemental sulphur and meet the following specifications:

UNIT	12 UNIT	22 UNIT
Total sulphur content, wt%	0.015	0.059
Doctor test	NEG	NEG
Flash point, °C, min.	45	45
Smoke point, mm, min.	24	24
Aromatics, vol% max.	20	20
Silver corrosion, max.	No. 1	No. 1

The TENDER DOCUMENTS request that the straight run kerosene also be hydrodesulfurised to a very low level of sulfur, namely 10 wt ppm, but because of the typical specifications worldwide for kerosenes we have taken this figure to be the maximum desired level of mercaptan sulfur in the product. For total sulfur, we have designed the UNITS to achieve a desulfurization rate of 90% and to thus reduce the sulfur in the N-RUMAILA kerosene to 0.015 wt% and in the MISCHRIF kerosene to 0.059 wt% total sulfur.

4.7.2 Description of UNIT

The process flow is illustrated in the attached simplified flow diagram; Dwg. No. 12-G001-D.

The charge stock, together with the make-up hydrogen and recycle gas, are heated to reaction temperature in exchanger and a fired heater.
In the reactor the contained sulfur, nitrogen, halogen and oxygen impurities, together with olefinic hydrocarbons, are converted over hydrobon catalysts. All the above reactions are exothermic, although the exothermic heat is relatively small when minimal quantities of these impurities are present. The reaction product is cooled through exchangers and a cooler en route to the product separator. Net separator gas is recycled and combined with hydrogen rich make-up gas while the separator liquid, after heat exchanged, is stripped in the fractionator column to remove H_2S and undesired light ends. The fractionator column bottoms product is sent to storage after heat exchange and cooling.

The H_2S rich sour gas is sent to the GAS/AMINE UNIT low pressure gas scrubber, while the undesired light ends is returned to the CRUDE UNIT.

When the chargestock is fed to this UNIT from the off-site tankage, this is stripped in the O_2 stripper column to remove oxygen for preventing serious plugging and corrosion in the UNIT.

4.7.3 Features of process

(1) Catalysts can be regenerated using a steam-air procedure.

(2) In consideration of cold charge from the off-site tankage, an O_2 stripper will be provided which precludes gummy matter to be created by dissolved oxygen and thus prevents reactors, columns, etc. from plugging and corrosion. Further, the O_2 stripper also serves as a feed surge drum, this enabling stabilized operation.

4.7.4 Estimated yield and properties

(1) N-RUMAILA kerosene

Chargestock: 11,000 BPSD 175 - 250°C TBP kerosene

Properties : 46.7 °API, 0.15 wt% sulphur, 14.5 vol% aromatics

	Wt%	SG	Vol% (SCFB)	BPSD
Feedstock:				
Raw kerosene	100.00	0.794	100.00	11,000
Hydrogen (Chemical)	0.12		(61)	
Product:				
Hydrogen sulfide	0.14			
Methane and ethane	0.01			
Propane produced	0.01			
Butanes produced	0.01	0.579	0.01	1
Pentane produced	0.02	0.629	0.03	3
Hexanes and all heavier material	99.93	0.791	100.26	11,029
Total:	100.12	-	100.30	11,033

C_6^+ product: 47.3 °API, 0.015 wt% sulfur, 14 vol% aromatics

(2) MISCHRIF kerosene

Chargestock: 17,500 BPSD 175 - 250°C TBP kerosene

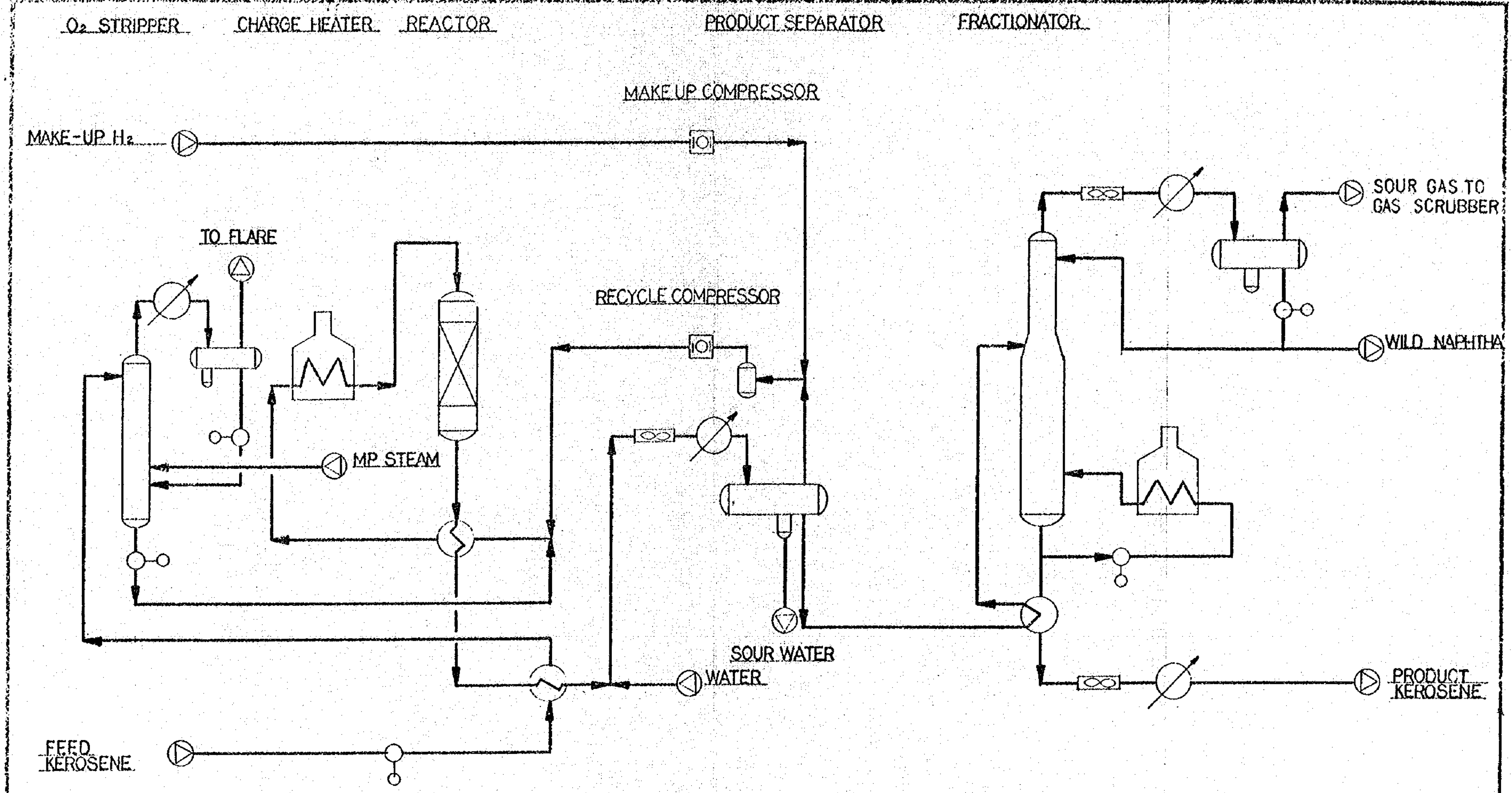
Properties : 44.7 °API, 0.593 wt% sulphur, 19.4 vol% aromatics

	Wt%	S.G.	Vol% (SCFB)	BPSD
Feedstock:				
Raw kerosene	100.00	0.803	100.00	17,500
Hydrogen (Chemical)	0.17		(91)	
Product:				
Hydrogen sulphide	0.57			
Methane and ethane	0.02			
Propane produced	0.02			
Butanes produced	0.03	0.579	0.04	7
Pentane produced	0.03	0.629	0.04	7
Hexanes and all heavier material	99.50	0.797	100.30	17,553
Total:	100.17	-	100.38	17,567

C_6^+ product: 46.1 °API, 0.059 wt% sulfur, 19 vol% aromatics

4.7.7 Battery limit conditions

Fluid	Temperature (°C)	Pressure (kg/cm2G)	Note
Undesulfurized kerosene	88	5.0	Hot charge
	10 - 40	5.0	Cold charge
Desulfurized kerosene	40	3.0	
Make-up gas	40	14.0	
Wild naphtha	40	3.0	
Sour gas	40	6.0	
Foul water	40	3.0	



JAPAN INTERNATIONAL
COOPERATION AGENCY

EXPORT REFINERY PROJECT

12/22
KEROSENE DESULFURIZATION UNIT
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 12-G-001

4.8 GO-HDS UNIT

4.8.1 Process design basis

(1) Capacity and charge

The unit processes all the MISCHRIF light and heavy gas oils and a portion of the N-RUMAILA heavy gas oil as produced from the CRUDE UNITS. It also is capable of treating the MISCHRIF gas oil only.

Capacity and chargestock of these UNITS are as follows:

	UNIT No. 13	UNIT No. 23
Crude source	N-RUMAILA	MISCHRIF
TBP cut point, °C	250 - 343	250 - 343
Capacity, m^3/hr	102	162
(BPSD)	(15,400)	(24,500)

UNIT No. 13 is a future unit. There will be cases where UNIT No. 23 will handle a part of the mixed N-RUMAILA HGO. However, the design is based on the case of treating the above MISCHRIF gas oil.

(2) Product specification

To hydrotreat the feedstock using H_2 rich gas from the CAT-REFORMERS. To stabilize the treated gas oils and the fractionate that are given the following fractions:

	TBP Cut Points, °C
Off gas	-
Naphtha and lighter	-
Desulfurized light gas oil	IBP - 300
Desulfurized heavy gas oil	300 - EP

Both desulfurized products to be free from H_2S and elemental sulfur and meet the following specifications after fractionation:

	Light Gas Oil	Heavy Gas Oil
Flash point, °C, min.	70	80
Copper strip corrosion, max.	1	1
Sulfur content, wt% max.	0.15	0.3
Acidity, mg. koh/g max.	1.0	1.0
Water, vol% max.	0.05	0.05
ASTM 5/95 gap. °C with HGO	To be stated	

Note: The units will be designed, as that the sulfur content of total gas oil will be 0.23 wt%.
The sulfur contents of LGO and HGO should be regarded as target values.

(3) Process design notes

These notes will conform to TENDER DOCUMENTS Vol. II. 10.1 and 10.8.5 in general, except for the following deviations:

Product stripping will be carried out by means of steam mainly with consideration to economy. In the case of the fractionator, steam will be injected into the bottom to prevent temperature increase at the bottom.

4.8.2 Description of UNIT

The process flow is illustrated in the attached simplified flow diagram; Dwg. No. 13-G001-D.

The chargestock, together with the make-up hydrogen and recycle gas, are heated to reaction temperature in exchangers and a fired heater.
In the reactor the contained sulfur, nitrogen, halogen and oxygen impurities, together with olefinic hydrocarbons, are converted over hydrobon catalysts, with metallic compounds being removed by a decomposition/adsorption mechanism. All the above reactions are exothermic, altough the exothermic heat is relatively small when minimal quantities of these impurities are present. The reaction product is cooled through exchangers and a cooler en route to the product separator.

Net separator gas is scrubbed with an amine solution, then this is recycled and combined with hydrogen rich make-up gas. Meanwhile the separator liquid, after heat exchange, is stripped in the stripper column to remove H_2S and undesired light ends. The H_2S and light ends are intoduced into the GAS/AMINE UNIT low pressure scrubber.
The stripper column bottoms product is heated through a heat exchanger and a heater, entering into the fractionator where it is fractionated to the LGO and the HGO fraction. The hydrotreated LGO, withdrawn as side cut product from the LGO stripper, is sent to storage after heat exchange and cooling.

The HGO is withdrawn from bottoms of the fractionator, heat exchanged and cooled. Then the HGO is introduced into the sand filter and the salt dryer, where it is dehydrated by salt and is sent to storage.

When the chargestock is fed to this UNIT from the off-site tankage, this is stripped in the O_2 stripper column to remove oxygen for preventing serious plugging and corrosion in this UNIT.

4.8.3 Features of process

(1) The catalysts can be regenerated using a steam-air procedure.

(2) For the desulfurization of the gas oils we have set the sulfur of the total product to be 0.23 wt% maximum.
For the FUTURE UNIT for N-RUMAILA gas oil this represents a desulfurization rate of about 82% and yield figures for this (in a UNIT similar to that described below for the MISCHRIF gas oil) are presented in Para. 4.8.4 (1).

The MISCHRIF gas oil needs a desulfurization rate of 90.8% and despite the increased chargestock quantity, this UNIT also needs to be equipped with only a single reactor provided the hydrogen rich recycle gas is scrubbed with a slip stream of amine solution from the sour gas treating facility.
With amine scrubbing and the use of a newly developed and highly efficient Hydrobon catalyst for heavy distillate desulfurization, the inventory, consumption and cost of catalyst per day can be truly minimised.

(3) As steam injection into the fractionator bottom is being considered, the following considerations have been given to the LGO and HGO to avoid the mixture of water with hydrotreated LGO and HGO:

While the LGO will be withdrawn as a fractionator side cut, the side cut stripper will be of the vacuum type for reducing water content. The HGO will be withdrawn from the fractionator bottom and cooled so as to reduce the water content via a sand filter and a salt dryer.

(4) In consideration of cold charge from the off-site tankage, an O_2 stripper will be installed to prevent gum generation due to dissolved oxygen, and plugging and corrosion at the reactor, columns, etc. Concurrently, the O_2 stripper will also has the function of a feed surge drum, thus resulting in stable operation.

4.8.4 Estimated yield and properties

(1) N-RUMAILA gas oil (Future unit)

Chargestock: 15,400 BPSD 250 - 343°C TBP gas oil

Properties : 36.1 °API, 1.25 wt% sulfur

	Wt%	SG	Vol% (SCFB)	BPSD
Feedstock:				
Raw oil charged	100.00	0.844	100.00	15,400
Hydrogen (Chemical)	0.28		(156)	
Product:				
Hydrogen sulfide	1.08			
Methane and ethane	0.04			
Propane produced	0.04			
Butanes produced	0.05	0.579	0.07	11
Pentane produced	0.07	0.629	0.09	14
Hexanes and all heavier material	99.00	0.832	100.48	15,474
Total:	100.28	-	100.64	15,499

C_6^+ product: 38.6 °API, 0.23 wt% sulfur

(2) MISCHRIF gas oil

Chargestock: 24,500 BPSD 250 - 343°C TBP gas oil

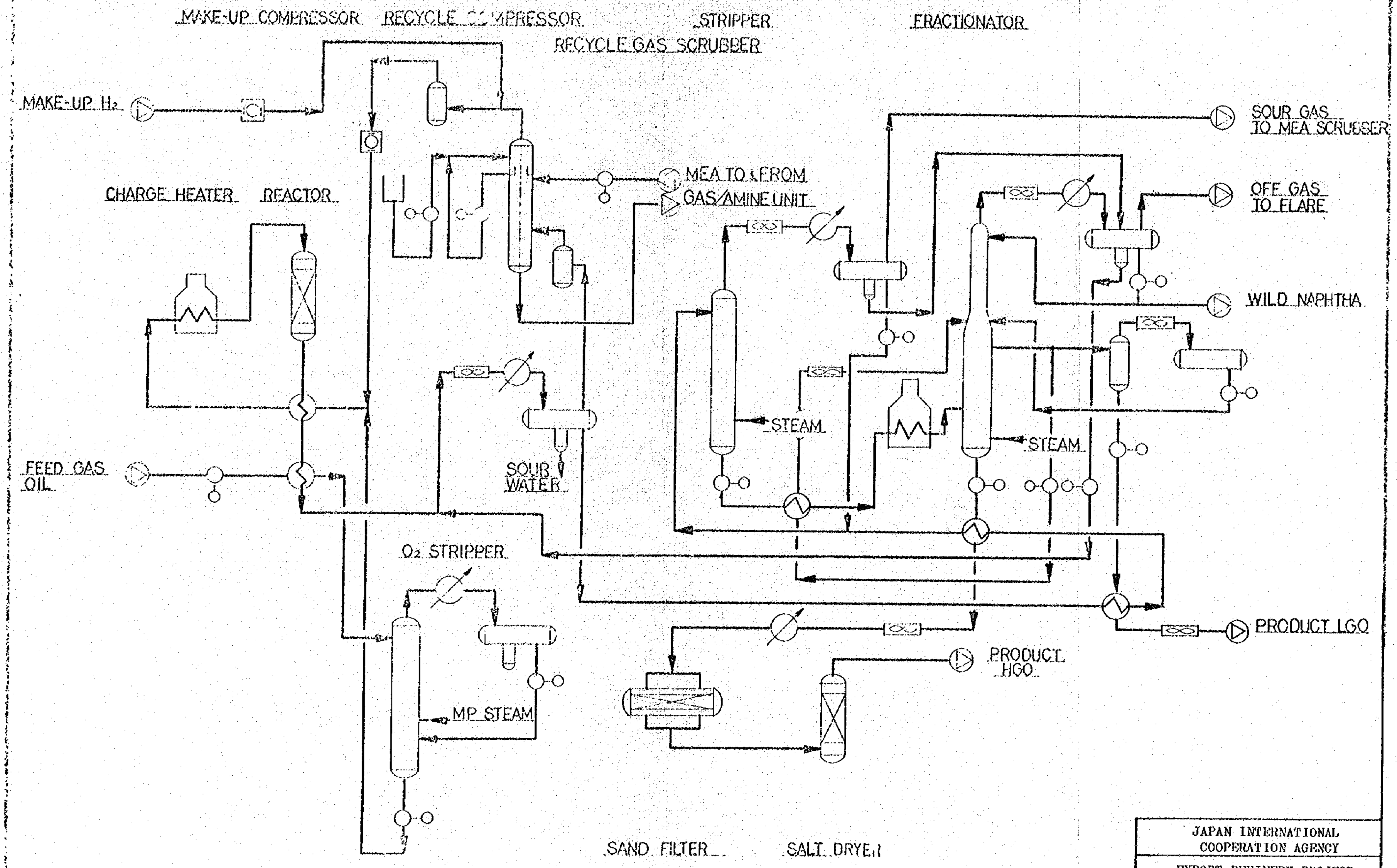
Properties : 32.4 °API, 2.50 wt% sulfur

	Wt%	SG	Vol% (SCFB)	BPSD
Feedstock:				
Raw oil charged	100.00	0.863	100.00	24,500
Hydrogen (Chemical)	0.45		(257)	
Product:				
Hydrogen sulfide	2.41			
Methane and ethane	0.09			
Propane produced	0.08			
Butanes produced	0.11	0.579	0.16	39
Pentane produced	0.14	0.629	0.19	46
Hexane and all heavier material	97.62	0.838	100.59	24,645
Total:	100.45	-	100.94	24,730

C_6^+ product: 37.4 °API, 0.23 wt% sulfur

4.8.7 Battery limit conditions

	Temperature (°C)	Pressure (kg/cm^2G)	Note
Undesulfurized gas oil	100	5.0	Hot charge
	10 - 40	5.0	Cold charge
Desulfurized HGO	65	3.0	
Desulfurized LGO	55	3.0	
Make-up naphtha	40	14.0	
Wild naphtha	40	3.0	
Sour gas	40	6.0	
Foul water	40	3.0	
Lean amine	40	7.0	
Rich amine	53	7.0	



JAPAN INTERNATIONAL
COOPERATION AGENCY

EXPORT REFINERY PROJECT

13/23
GAS OIL DESULFURIZATION UNIT
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 13-G-001

4.9 GULF HDS UNIT

4.9.1 Process design basis

(1) Design capacity

Design capacities are as follows:

	Unit No. 14	Unit No. 24
Crude source	N-RUMAILA	MISCHRIF
Capacity, m^3/hr	596	497
(BPSD)	90,000	75,000

(2) Feed specifications

Residual feed oil specifications for the proposal use are as follows:

Crude source	N-RUMAILA	MISCHRIF
Nominal TBP cut point, °C	343	343
API gravity	16.2	11.1
Sulfur, wt%	3.4	6.05
Nitrogen, wt ppm	2,500	2,280
Carbon residue, Rams, wt%	10.2	11.0
Nickel, wt ppm	12	17
Vanadium, wt ppm	50	112
Iron, wt ppm	3	1
Viscosity, CS at 93.9°C	26	55
CS at 50°C	220	900
Pour point, °C	15	12.8
Asphaltens, wt%	2.70	2.35
Salt as Nacl, ptb	1.0 max.	1.0 max.

(3) Product specifications

Producing the following fractions in normal operation.

(a) Petrochemical naphtha

Specific gravity, 15/4°C max. : 0.730

ASTM Distillation, °C

Initial boiling point, min. : 30
Final boiling point, max. : 171

Sulfur, wt%, max. : 0.01

Olefins, vol%, max. : 1.0

(b) 0.3% sulfur fuel oil

Sulfur, wt%, max. : 0.3

Flash point, °C, min. : 65.6

Viscosity, CS at 37.8°C : 32 - 65

Pour point, °C, max. : 21.1

Ash, wt%, max. : 0.10

Water & sediment, wt%, max. : 0.5

Conradson carbon res., wt%, max.: 12

(c) 1.0% sulfur fuel oil

Sulfur, wt%, max. : 1.0

Flash point, °C, max. : 65.6

Viscosity, CS at 37.8°C : 75 - 162

Pour point, °C, max. : 21.1

Ash, wt%, max. : 0.10

Water & sediment, wt%, max. : 0.5

Conradson carbon res., wt%, max.: 12

(4) Process design notes

(a) Conceptual process design was made, using the proposal data, which have been provided by Gulf Science and Technology Company.

(b) Design capacities have been determined with the following consideration points.

(i) The home fuel oil will be split from both N-RUMAILA and MISCHRIF atmospheric residues. The total fuel oil consumption is 18,600 BPSD.

(ii) Allowance should be considered on the MISCHRIF HDS UNIT since the crude oil characteristics are not clear at present, while the N-RUMAILA crude oil has similar characteristics with the other Middle East crude oil experienced in the commercial HDS UNITS in Japan. The MISCHRIF HDS UNIT allows 10.5 percent excess throughout capacity.

(c) The semiannual (or nominal "Six-month") catalyst cycle, which is the based of the proposed design, can be fitted to either a one-year or two-year turn around schedule. No catalyst regeneration will be done.

(d) Following ancillary process sections are served as common facilities of the GULF HDS UNITS:

(i) LP off gas sweetening

(ii) HDS amine regeneration

(iii) Waste water stripping

4.9.2 Description of unit

The following process description of the unit refers to the Block Flow Diagram for the UNIT, Dwg. No. 14/24-G-001.

(1) High pressure reactor section:

The process is based upon a two-stage (Type III) catalytic fixed bed, high pressure, hydrodesulfurization technology. In this section, the residual feed is combined with hydrogen-rich gas, then the mixture is sent to the reactor where desulfurization, denitrification and demetallation reactions take place. The reactions consume hydrogen and produce hydrogen sulfide, ammonia and light hydrocarbons.

(2) Recycle gas treating section:

Flashed gas separated in the high pressure reactor section is sent, to the recycle gas treating section to increase the hydrogen purity and to remove the bulk of hydrogen sulfide produced.

(3) Fractionation section:

Flashed liquid separated in the high pressure reactor section is sent to the fractionation section where the liquid is split into three streams: a residual fuel oil as a fractionator bottoms product, a stabilized naphtha as a stabilizer bottoms product and a sour off gas as an overhead gas product. The residual fuel oil and the stabilized naphtha are sent to storage as final products.

(4) LP off gas sweetening section:

The off gas from the fractionation section is amine-scrubbed and is sent to refinery fuel gas system.

(5) HDS amine regeneration section:

This section has a function to regenerate amine solution which is exclusively used in the GULF HDS UNITS. Because of large amine capacity, two identical trains are employed. A conventional regeneration technology is applied and wet overhead vapors are sent to the SR-UNIT.

(6) Waste water stripping section:

Waste water streams are assembled and sent to a waste water stripper where ammonia and hydrogen sulfide are stripped. The wet ammonia-bearing acid gas as an overhead product is sent to the SR-UNIT. A part of stripped water is reused for recycle gas wash and a balance is transferred to the CRUDE UNIT as an injection water of crude oil desalting.

4.9.3 Estimated yields and propertes

(1) Yield

Crude source	N-RUMAILA	MISCHRIF
Naphtha, vol% charge	3.4	5.7
Fuel oil, vol% charge	98.4	97.2

(2) Properties

(a) Naphtha

Crude source	N-RUMAILA	MISCHRIF
Specific gravity, 60/60°F	0.7788	0.7788
Sulfur, wt%	0.03	0.02
Nitrogen, wt ppm	150	150
ASTM distillation, °C		
IBP	66	66
FBP	199	199
Olefins, vol%	2	2

(b) Desulfurized fuel oil (i)

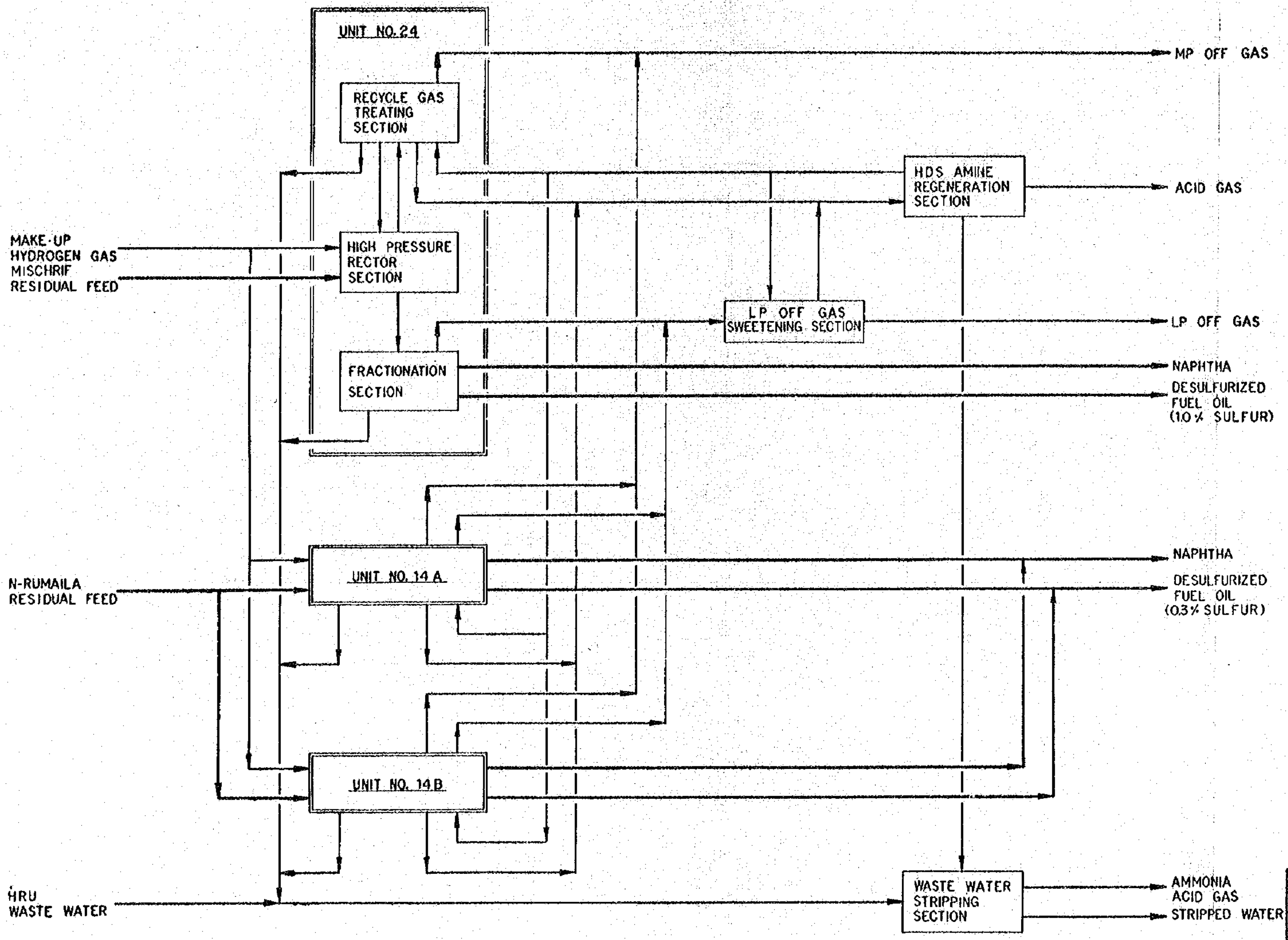
Crude source	N-RUMAILA	MISCHRIF
API gravity	22.2/24.2	20.4/24.9
Flash point, °C	70 (min.)	70 (min.)
Ash, wt%	0.004	0.008
Pour point, °C	16/0	21/8
Thermal stability	No. 1	No. 1
Viscosity, CS at 38°C	86/43	108/76
Sulfur, wt%	0.3	1.0

Conradson carbon, wt%	3.8/4.0	5.2/5.6
Nickel, wt ppm	1.0/1.5	4.5/5.0
Vanadium, wt ppm	2.5/3.0	13.5/14.5

Note (i) Ranges indicate OSR/EOR variations.

4.9.4 Battery limit conditions

Fluid	Temperature (°C)	Pressure (kg/cm^{2}G)	Note
Residual feed	50 - 70	-0.4	From tankage
Naphtha product	38	5.0	Rundown
Desulfurized fuel oil product	70	5.0	Rundown
Make-up hydrogen gas	40	20	From HYDROGEN UNIT
LP off gas			To fuel gas system
Stripped water	52	5.0	To CRUDE UNIT
MP off gas			To HYDROGEN UNIT
Acid gas	38	0.7	To SR-UNIT
Ammonia acid gas	85	1.0	To SR-UNIT



JAPAN INTERNATIONAL
COOPERATION AGENCY

EXPORT REFINERY PROJECT

GULF HDS UNIT
BLOCK FLOW DIAGRAM

DWG. NO. 14/24-G-001

4.10 HYDROGEN UNIT

4.10.1 Process design basis

(1) Design capacity

Design capacities are as follows:

	Unit No. 02A	Unit No. 02B
Capacity, Nm^3/day (as 100% H_2)	1,900,000	1,900,000

(2) Feed gas and boiler feed water specifications

(a) Feed gas specification

	mol%
Hydrogen	27.60
Methane	31.84
Ethane	16.60
Propane	15.00
Butanes	6.77
Pentanes	2.18
Hydrogen sulfide	50 ppm (max.)
Total:	100.00
Pressure, kg/cm^{2}G	3.5
Temperature, °C	40.0

(b) Boiler feed water specification

Acidity, pH range : 7.0 ÷ 9.0

Total hardness : Net detectable

Oil content, wt ppm : 0

Fe, wt ppm : 0.05 max.

SiO_2, wt ppm : 0.20 max.

Cu, wt ppm : 0.01 max.

(3) Product specification

Hydrogen, mol% : 90.0 min.

Methane, mol% : 3.0 max.

Carbon monoxide

Plus carbon dioxide, vol ppm: 10.0 max.
Water : Saturated

Pressure, kg/cm^2G : 20

Temperature, °C : 40

(4) Process design notes

(a) The feed gas is the separated hydrocarbon-rich gas leaving Hydrogen Recovery Unit and the refinery off-gas.

(b) 2-train lay out is adopted in accordance with Appendix I, page 28 of 62 and from the view point of flexibility and operational easiness.

(c) Design capacity is 2 x 1,900,000 Nm^3/day as 100% hydrogen. However, maximum hydrogen requirement by the GULF HDS UNITS and future Gas Oil Desulfurization Unit is approximately 3,400,000 Nm^3/day (AS 100% H_2).

This 10% excess design allows flexibility of HYDROGEN UNIT in case of deviated operation conditions from design, such as heavier or lighter feed gas and/or fuel gas.

(d) One-year continuous operation is considered as a design base of the HYDROGEN UNIT. However, expected catalyst life is 3 to 5 years.

4.10.2 Description of unit

The following process description of the unit refers to the Simplified Process Flow Diagram for the UNIT, Dwg. No. 02-G-001.

(1) Hydrogen recovery unit

This unit consists of three subsections, i.e. (1) Feed Gas Compression and Amine Absorption, (2) Amine Regeneration, and (3) Feed Gas Pretreatment and Cryogenic Hydrogen Separation. The cryogenic separation is done by multistage process and refrigeration is produced by Joule-Thomson Effect from isenthalpic expansion of hydrocarbons. The product hydrogen of 97 percent purity is obtained at 90 percent of recovery rate and is sent to the GULD HDS UNITS. The separated hydrocarbon rich gas is used as feed gas for hydrogen manufacturer.

(2) Final desulfurization and reforming section

Feed gas is fed to the final desulfurization section. Pressurized gas is then heated and sent to the hydrogenerator in which the organic sulfur is converted to H_2S, and the olefinic hydrocarbon to a parafinic hydrocarbon. The feed gas is then introduced into the H_2S adsorber where the H_2S is removed from the stream almost completely, using zinc oxide, in the form of ZnS.

The desulfurized gas is then mixed with the process steam and charged to the preheater coil located in the convection zone of the steam reformer and heated to the necessary temperature for the reforming reactions.

The steam reforming reactions of steam and hydrocarbons occurs in the presence of a nickel-based catalyst in the catalyst tubes, located in the radiant chamber of the steam reformer.

Effluent gas from the steam reformer consists of hydrogen, carbon monoxide, carbon dioxide, steam and some unconverted methane. This effluent gas is cooled through the No. 1 waste heat boiler which generates the process steam in the shell side.

Hot flue gas from the radiant chamber of the steam reformer is fully utilized to heat the mixture of feed gas and process steam, to generate the steam and to preheat the boiler feed water. Then, the flue gas is purged to the atmosphere.

(3) CO shift section

The effluent gas, from the No. 1 waste heat boiler, is introduced into the HTS converter in which the Fe_2O_3 catalyst is loaded. In this reactor, a large part of CO is converted to H_2 and CO_2 by reacting with excess steam. The shifted gas, then, passes through heat exchangers to cool and to recover the sensible heat of the process gas before entering the LTS converter in which the CuO-ZnO catalyst is loaded. After the two stage CO shift conversion, the CO in the process gas is almost converted to to H_2 and CO_2.

(4) CO_2 removal section

The shifted gas is cooled through the reboiler as solution boiling in the CO_2 removal section. Then the process condensate is separated and the shifted gas is introduced into the absorber. In the absorber, almost all of the CO_2 gas is absorbed in the catacarb solution, and the effluent gas, from the absorber is further scrubbed in the mist separator, from a water spray to remove the entrained mist of the CO_2 wash solution.

The CO_2 rich solution, from the absorber, is sent to the regenerator in order to strip the absorbed CO_2 and to regenerate the catacarb solution. Regeneration of the solution is done by a heat supply in the reboiler.

(5) Methanation section

Effluent gas, from the absorber is heated from the heat exchanger with the methanator effluent and HTS converter effluent. Then, led to the methanator where the remaining CO and CO_2 is converted into a CH_4 in the presence of nickel-based methanation catalyst. The methanated gas is heat exchanged with the methanator feed stream, cooled through coolers. Then the process condensate is separated, again.

The process condensate, from separators is introduced to the degasser in order to strip the dissolved gases; then, cooled and sent to a process condensate treating section.

(6) Steam section

The boiler feed water, from the process condensate treating section, is preheated, and mixed with steam condensate from the GULF HDS UNIT; then, introduced to the deaerator, in which dissolved oxygen is stripped. The deaerated boiler feed water is pumped up to the steam drum.

The boiler feed water is sent to the steam generation coil, the No. 1 and No. 2 waste heat boiler in which the HP steam is generated.

The process steam, from the steam drum is mixed with the process hydrocarbon, and the rest of the steam is utilized as a motive steam in steam turbines.

A part of the boiler feed water is sent to the LP steam generator to generate the LP steam.

(7) Process condensate treating section

This section has a function to reuse process condensate. Process condensate is passed through the activated carbon to remove the COD and other organic matter and is polished to the degree of quality required in passage to the boiler by going through the mixed bed of anion and cation exchange resins. Then, it is stored in the demineralized water tank and sent to the HYDROGEN UNITS. The spent carbon is reactivated by burning off and the regeneration of the resin used with sulfuric acid as the cation resin and caustic soda as the anion resin, as a repeated use. The regeneration waste water, after adjusting the pH, is pumped out.

4.10.3 Features of unit

(1) Steam reformer

(a) Catalyst tube arrangement is one-line chamber type with radiant wall burners, thus high heat flux and uniform heating operating is easily achieved.

(b) The hair-pin shaped tubes are reliable and recommendable for feed inlet connection of the catalyst tubes.

(c) The following points should be taken into consideration comprehensively.

- o Allowable stress for high alloyed materials
- o Expected stress
- o Expected metal temperature
- o Thermal expansion
- o Dimensions
- o Selection of high alloy materials

(2) CO shift conversion

2 stage CO shift conversion, namely HTS conversion and LTS conversion is used. Since the LTS catalyst is sensitive to sulfur poisoning and high temperature, the catalyst lacks in its reliability. However, these problems are now entirely resolved by our technical competence and, as a result, the life of the LTS catalyst is presently extended to more than 4 years.

(3) CO_2 removal

Catacarb Process, which is one of hot potash wash processes, is selected because of its lower investment cost, lower regeneration heat requirements, and relatively low solution losses, compared to amine processes.

Because of the corrosive nature of hot potash solution, selection of proper materials, use of proved corrosion inhibitors, and special considerations for the design of tower internal structures are necessary.

Further, the foaming tendency of this solution makes it necessary to install a filter for the removal of small particles in the solution and to use suitable anit-foaming chemicals.

(4) Methanation

An individual "Fail-safe" automatic shut down system is provided to prevent temperature run-away in the methanator catalyst which will be caused by operational upset in the CO_2 removal section.

4.10.4 Material balance

(Design base/unit)

(1) Feed gas	kg·mol/hr
H_2	202.1
CH_4	233.2
C_2H_6	121.6
C_3H_8	109.9
C_4H_{10}	49.6
C_5H_{12}	16.0
Total:	732.4
H_2S, vol ppm	50.0 (maximum)
Pressure, kg/cm^{2}G	3.5
Temperature, °C	40.0

(2) Product hydrogen	kg·mol/hr
H_2	3,532.0
CH_4	109.4
H_2O	13.3
Total:	3,654.5
$CO + CO_2$, vol ppm	10.0 (maximum)
Pressure, kg/cm^{2}G	20.0
Temperature, °C	40.0

(Design base/unit)

(3) CO_2 gas

	kg·mol/hr
H_2	8.9
CO	0.02
CO_2	969.1
N_2	1.1
CH_4	0.2
H_2O	425.0
Total:	1,404.3
Pressure, kg/cm^{2}G	0.3
Temperature, °C	76.0

(4) HP steam

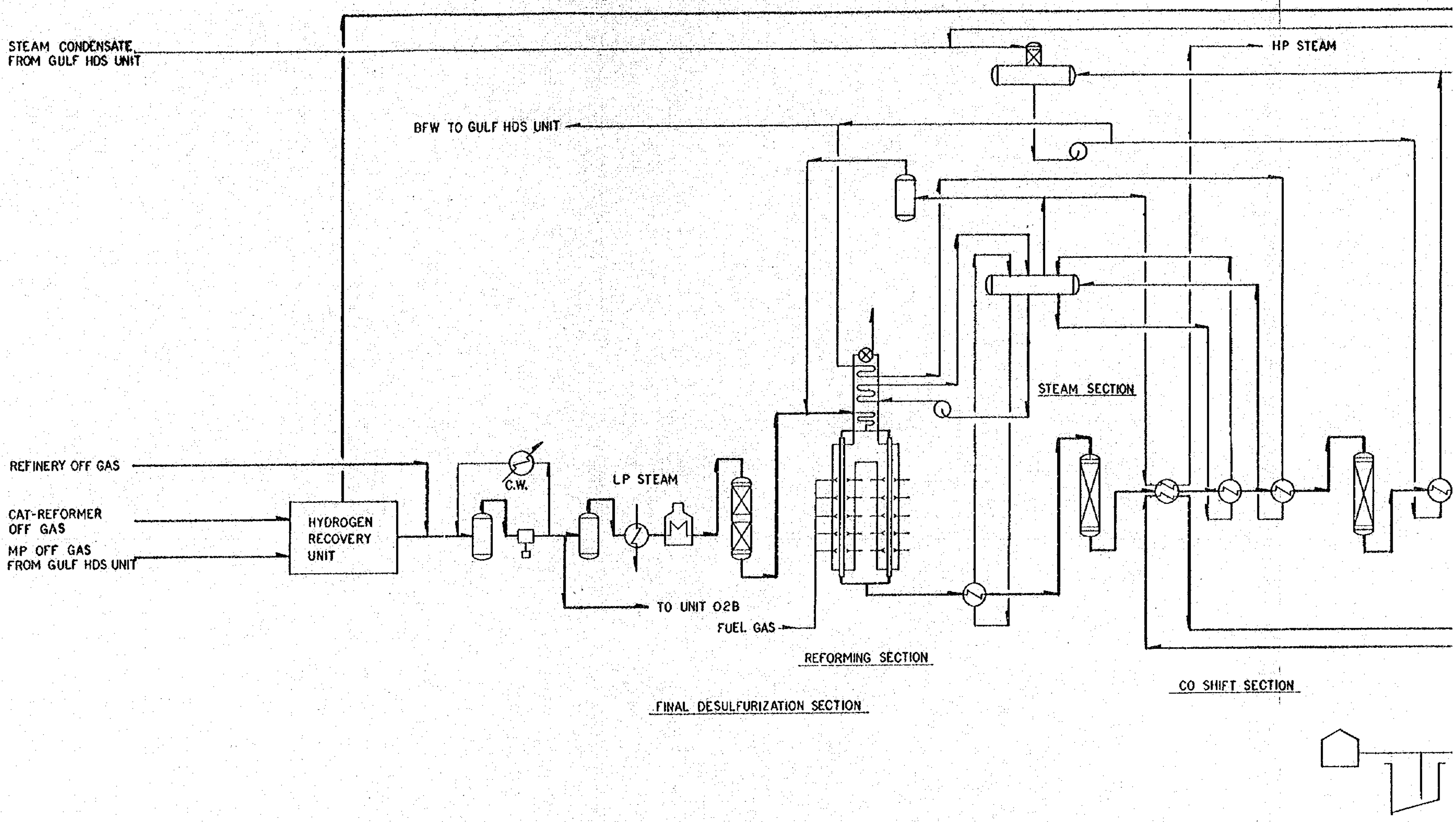
	kg·mol/hr
HP steam, kg/hr	27,950.0
Pressure, kg/cm^{2}G	32.0
Temperature, °C	350.0

(5) LP steam

	kg·mol/hr
LP steam, kg/hr	26,232.0
Pressure, kg/cm^{2}G	3.5
Temperature, °C	147.0

4.10.5 Battery limit conditions

Fluid	Temperature (°C)	Pressure (kg/cm2G)	Note
Feed gas	40	3.5	
Product hydrogen	40	20	
HP steam	350	32	
LP steam	147	3.5	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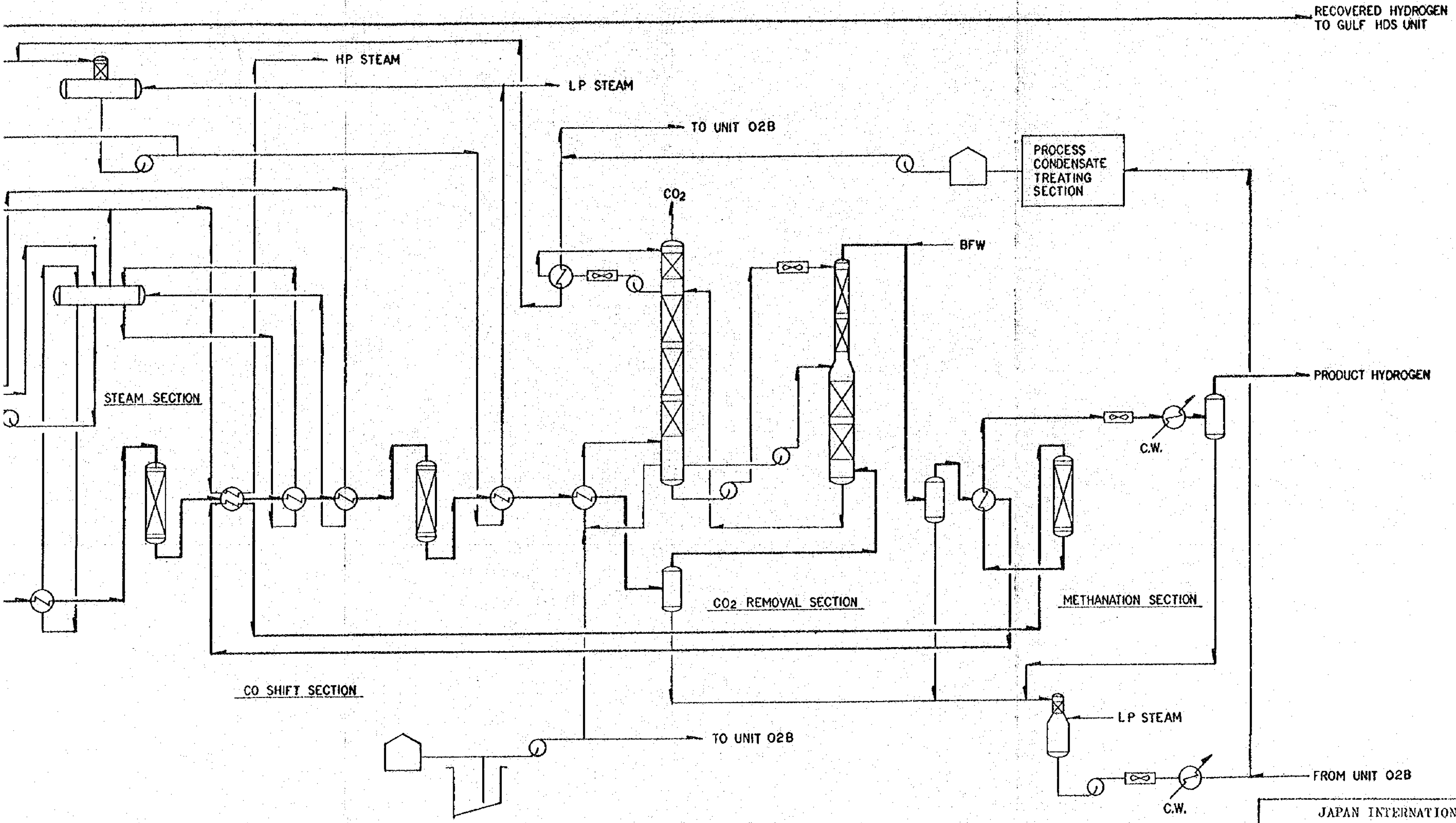
HP STEAM
BFW TO GULF HDS UNIT
STEAM SECTION
REFINERY OFF GAS
CAT-REFORMER OFF GAS
MP OFF GAS FROM GULF HDS UNIT
HYDROGEN RECOVERY UNIT
C.W.
LP STEAM
TO UNIT 02B
FUEL GAS
REFORMING SECTION
CO SHIFT SECTION
FINAL DESULFURIZATION SECTION



JAPAN INTERNATIONAL
COOPERATION AGENCY

EXPORT REFINERY PROJECT

HYDROGEN UNIT
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 02-G-001

4.11 GAS/AMINE UNIT

4.11.1 Process design basis

(1) Capacity and charge

This UNIT is intended to treat the streams from the following UNITS:

Sour gas : N-HDS UNIT stabilizer overhead off-gas
K-HDS UNIT fractionator overhead off-gas
GO-HDS UNIT stripper overhead off-gas

Sour LGP : N-HDS UNIT stabilizer overhead liquied product

Sweet LPG: CAT-REFORMER debutanizer overhead liquid product

The capacity of the UNIT will be sufficient to treat 8,465 Nm^3/hr (except for the FUTURE UNIT) of sour gas and 69 m^3/hr of sour LPG.

Also, the UNIT will be capable of separating 83 m^3/hr of sweet LPG into C_4 LPG and C_3 lighter.

Amine solution used to treat GO-HDS UNIT recycle gas will be regenerated in the amine regenerator.

Note: The UNIT intended to treat the sour gas streams from the GULF HDS UNIT will be designed together with the GULF HDS COMPLEX.

(2) Product specifications

The quality of products from this unit will be as shown below.

(a) Sweet fuel gas

0.16 g/Nm^3 of H_2S on a dry gas basis

(b) C_3 LPG

Total sulfur, wt ppm: 100

(c) C_4 LPG

Pentanes and heavier: 2.0 LV% max.
Propane and lighter : 2.0 LV% max.
Total sulfur, wt ppm: 100

(d) Acid gas

H_2S content : 97.0 vol% (dry basis)

(3) Turn down

The UNIT will be capable of 30% reduced operation.

(4) Process design notes

The process design notes given in TENDER DOCUMENTS 10.1 and 10.12.5 will be satisfied.

4.11.2 Description of UNIT

The process flow is illustrated in the attached simplified flow diagram Dwg. No. 01-G001-D.
This/UNIT will consist of the amine treating section and the gas recovery section.

The amine treating section will consist of the low pressure gas scrubber, LPG extractor and regenerator.

Low pressure sour gas from each HDS UNIT will be led to the low pressure gas scrubber where H_2S contained in the gas will be removed by absorption into MEA solution, and the sweet gas will be sent to the fuel gas facility in the off-site area.

Sour LPG withdrawn as liquid product from the N-HDS UNIT stabilizer overhead receiver will enter the LPG extractor where H_2S will be removed by absorption into MEA solution, and the sweet LPG will be sent to the gas recovery section.

After absorption of H_2S in the LPG extractor, the rich MEA solution will join in the rich MEA solution stream from the GO-HDS UNIT recycle gas scrubber, and the stream will be led to the amine flash drum.

After flashing of light hydrocarbons in the amine flash drum, the MEA solution stream will join in the rich MEA solution stream from the low pressure gas scrubber, and after heat exchange, the joint stream will pass to the MEA regenerator.

The regenerator will be provided with a reboiler heated by LP steam. In the regenerator, the rich MEA solution will be regenerated and withdrawn as lean MEA solution from the bottom section.

Column overhead vapor containing acid gas released from the rich MEA solution will be cooled to 40°C in the air condenser and in the water condenser and will enter the overhead receiver, where it will be separated into acid gas and reflux, and the acid gas will be sent to the SR-UNIT.

After heat exchange, the lean solution will enter the amine solution tank and will be cooled to a required temperature in the air cooler and/or water cooler. Being pumped up to a required pressure, the lean solution will be sent to the LPG extractor, low pressure gas scrubber and GO-HDS recycle gas scrubber.

LPG free of H_2S after treatment in the LPG extractor will pass to the coalescer where entrained MEA will be removed, and the LPG will enter the LPG surge drum.

Also, LPG withdrawn as platformer debutanizer overhead liquid will be led to the LPG surge drum, and being pumped up to a required pressure, after heat exchange, the LPG will be fed to the depropanizer.

Depropanizer overhead vapor will be cooled to 40°C in the water condenser, where it will be totally condensed and passed to the overhead receiver.

Part of the condensed product will be withdrawn as C_3 and lighter product and the rest of it will be returned as a reflux to the tower.

C_4 LPG withdrawn from the bottom of the tower, after heat exchange, will be cooled to 40°C in the water cooler and will be sent to storage.

4.11.3 Features of process

(1) Since 15 wt% MEA solution has been selected, operation will be stable and easy and the UNIT has a flexibility for fluctuations in the feed.

(2) To enable 30% reduced operation, the heat exchangers, pumps and trays have been carefully designed.

(3) The temperature of the lean MEA solution to the LP gas scrubber has been raised higher than that of the gas, to prevent heavier components in the gas from condensing.

(4) Since lean MEA solution is passed to the amine tank and is sent to the scrubber and to the extractor, the stability and flexibility of operation is larger.

(5) The reclaimer has been provided.

(6) The UNIT has been designed to be capable of treating recycle gas and sour gas from the future GO-HDS UNIT in addition to those from the UNITS to be constructed in the present project.

4.11.4 Estimated yields and properties

The estimated yields and properties are as shown below.

(1) Amine treating section

Mol%	LP Gas Scrubber		LPG Extractor	
	LP Sour Gas	LP Sweet Gas	Sour LPG	Sweet LPG
H_2S	18.3	(a)	1.5	(b)
H_2	42.6	52.2	-	-
C_1	9.2	11.3	-	-
C_2	15.8	19.3	5.5	5.6
C_3	8.7	10.7	36.0	36.6
C_4	4.0	4.8	56.8	57.6
C_5	1.4	1.7	0.2	0.2
C_6^+	Nil	Nil	Nil	Nil
Total	100.0	100.0	100.0	100.0
MW	20.4	17.4	51.1	51.3
Nm^3/hr	8,465(c)	6,920(c)	-	-
m^3/hr @15°C	-	-	69	68.5

Note: (a) 0.16 g/Nm^3

(b) 100 wt ppm

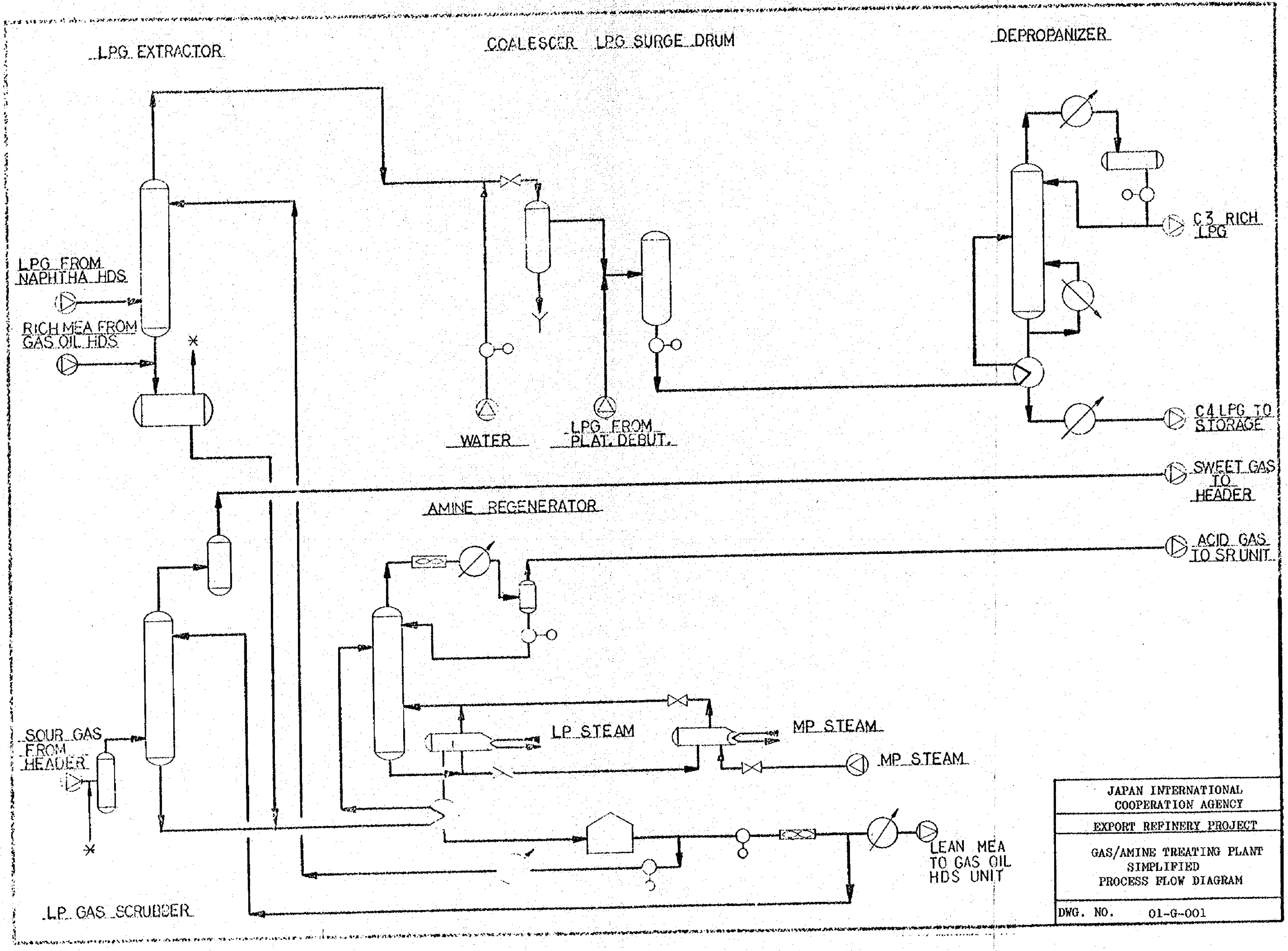
(c) Excluding that from the future GO-HDS UNIT

(2) Gas recovery section

Mol%	LPG from Extractor	LPG from Reformer	Total LPG to Depropanizer	C_3 and Lighters	C_4 LPG
H_2	-	0.4	Nil	0.1	-
C_1	-	0.7	0.1	0.3	-
C_2	5.6	8.6	6.1	14.0	-
C_3	36.6	33.2	36.0	82.7	Nil
C_4	57.6	56.7	57.5	2.9	99.6
C_5	0.2	0.4	0.2	-	0.4
C_6^+	Nil	Nil	Nil	-	Nil
Total	100.0	100.0	100.0	100.0	100.0
MW	51.3	50.5	51.2	42.3	58.0
m^3/hr @15°C	68.5	14.5	83.0	32.5	50.5

4.11.7 Battery limit conditions

Fluid	Temperature (°C)	Pressure (kg/cm^2G)	Note
Sour LPG	40	15.0	
Sweet LPG	40	12.0	
C_4 LPG	40	17.0	
C_3 LPG	40	19.0	
Sour gas	40	6.0	
Sweet gas	60	3.88	
Acid gas	40	0.6	
Rich amine	53	7.0	From GO-HDS
Lean amine	40	7.0	For GO-HDS
Acid gas drain	40	3.0	



JAPAN INTERNATIONAL COOPERATION AGENCY

EXPORT REFINERY PROJECT

GAS/AMINE TREATING PLANT
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 01-G-001

4.12 SR-UNIT

4.12.1 Process design basis

(1) Plant capacity

600 tons/day 2 trains, based on H_2S in feed acid gas

(2) Feed gas conditions

Clean Acid Gas

Temperature : 38°C

Pressure : 0.45 kg/cm^{2}G

Compnents

H_2S : 97 vol% (Dry gas base)
HC : 3 vol% (Dry gas base)
H_2O : 0.7 kg/cm^{2}G, 38°C saturated
Flow rate: 34,075 Nm^3/hr (Dry gas base)

NH_3 Acid Gas

Temperature : 85°C

Pressure : 0.45 kg/cm^{2}G

Components

H_2S : 50 vol% (Dry gas base)
NH_3 : 50 vol% (Dry gas base)
H_2O : 1.2 kg/cm^{2}G, 85°C saturated
Flow rate: 3,784 Nm^3/hr (Dry gas base)

(3) Specification of sulfur recovery rate

95% minimum

(4) Operating range

20% - 100% of normal

4.12.2 Process description

(1) Principle of process

The basic sulfur recovery process is the one called the Claus process in which hydrogen sulfide is converted into sulfur by oxidation.
The reaction is formularized as follows:

$$H_2S + 1/2\ O_2 \longrightarrow H_2O + S \qquad (1)$$

However, in actual sulfur recovery units, the formation of elementary sulfur proceeds through the following two stages;

$$H_2S + 3/2\ O_2 \longrightarrow SO_2 + H_2O \qquad (2)$$

$$2H_2S + SO_2 \longrightarrow 2H_2O + S \qquad (3)$$

Namely, 1/3 of the hydrogen sulfide fed to the unit is converted into sulfur dioxide as shown in Formula (2), which reacts with the remaining hydrogen sulfide subsequently and yields sulfur.

The major high temperature reactions in the reaction chamber are those shown in Formula (2) and (3), and various side reactions follow.

The catalytic reaction in the reactors followed by the above is chiefly the one shown in Formula (3). NH_3 gas in the feed gas is burned in the reaction chamber into Nitrogen and steam as shown in Formula (4).

$$2NH_3 + 3/2\ O_2 \longrightarrow N_2 + 3H_2O \qquad (4)$$

(2) Process flow

The process flow is illustrated in the attached simplified flow diagram Dwg. No. 03-G001-D.

Acid gas from amine units and waste water treating units will come together and after being preheated with an acid gas preheater, will be charged to the main chamber acid gas K.O. drum.

The acid gas from the main chamber acid gas K.O. drum will be separated and fed to the A train sulfur recovery unit and B train sulfur recovery unit.
And the acid gas will be burned, with air controlled stoichiometrically, so that the ratio of H_2S to SO_2 in the sulfur coalescer outlet gas will be 2.

The reaction gas mixture from the reaction chamber will pass through the waste heat boiler and enter the No. 1 condenser in which it will be cooled to separate as much gaseous sulfur as possible from the reaction gas.

Before entering the No. 1 reactor, the reaction gas will be mixed with high temperature burning gas from the No. 1 inline burner. As a result of this, the temperature of the gas will rise, which will be maintained at the reactor inlet.

H_2S and SO_2 will react over the catalyst in the reactor, being converted into gaseous sulfur.

The reaction gas will be led from the No. 1 reactor to the No. 2 condenser where the reaction gas will be cooled, with the reaction heat being removed, and condensed liquid sulfur will flow to the sulfur pit.

The completeness of the catalyst reaction will be ensured and the recovery rate will be increased as much as possible by repeating the above procedure twice.

The last stage condenser effluent gas will be led to the sulfur coalescer, in which almost all sulfur mist will be removed.

Sulfur coalescer effluent gas will be fed to incinerator and sulfur compounds in its gas will be burnt to SO_2.
In the waste heat boiler, M.P. steam will be generated and it will be collected to the M.P. steam drum.

Likewise, in the condensers L.P. steam will be generated and it will be collected to the L.P. steam drum.

Liquid sulfur which is collected from the waste heat boiler and the condensers in the sulfur pit will be de-gassed with the H_2S stripper and discharged to the liquid sulfur tanks.

4.12.3 Features of process

As features of the present process, the following will be mentioned;

(1) While the TENDER DOCUMENTS specify that NH_3 acid gas emitted from H_2S-bearing water shall be burned in an incinerator and off-gas released into the atmosphere, in consideration of the considerable amounts of H_2S and NH_3 contained in the NH_3 acid gas, release of NH_3 acid gas into the atmosphere will not only cause sulfur loss but will give rise to air pollution.

Therefore, the present process contains a sulfur recovery unit for processing NH_3 acid gas and recovering sulfur.

(2) Due to the adoption of inline burner system for the purpose of controlling the reactor inlet temperature, operation is easy and also the sulfur recovery rate can be maintained at a high level.

(3) NH_3 contained in acid gas is also apt to cause plugging of plant facilities through reaction. For eliminating this trouble, preheater have been provided, together with the burners of special design. These enable stabilized continuous operation.

(4) For removing H_2S from liquid sulfur, a H_2S stripper has been provided to minimize H_2S content.

4.12.4 Estimated yield and properties

Yield : Liquid sulfur 1,140 T/D (Max.)

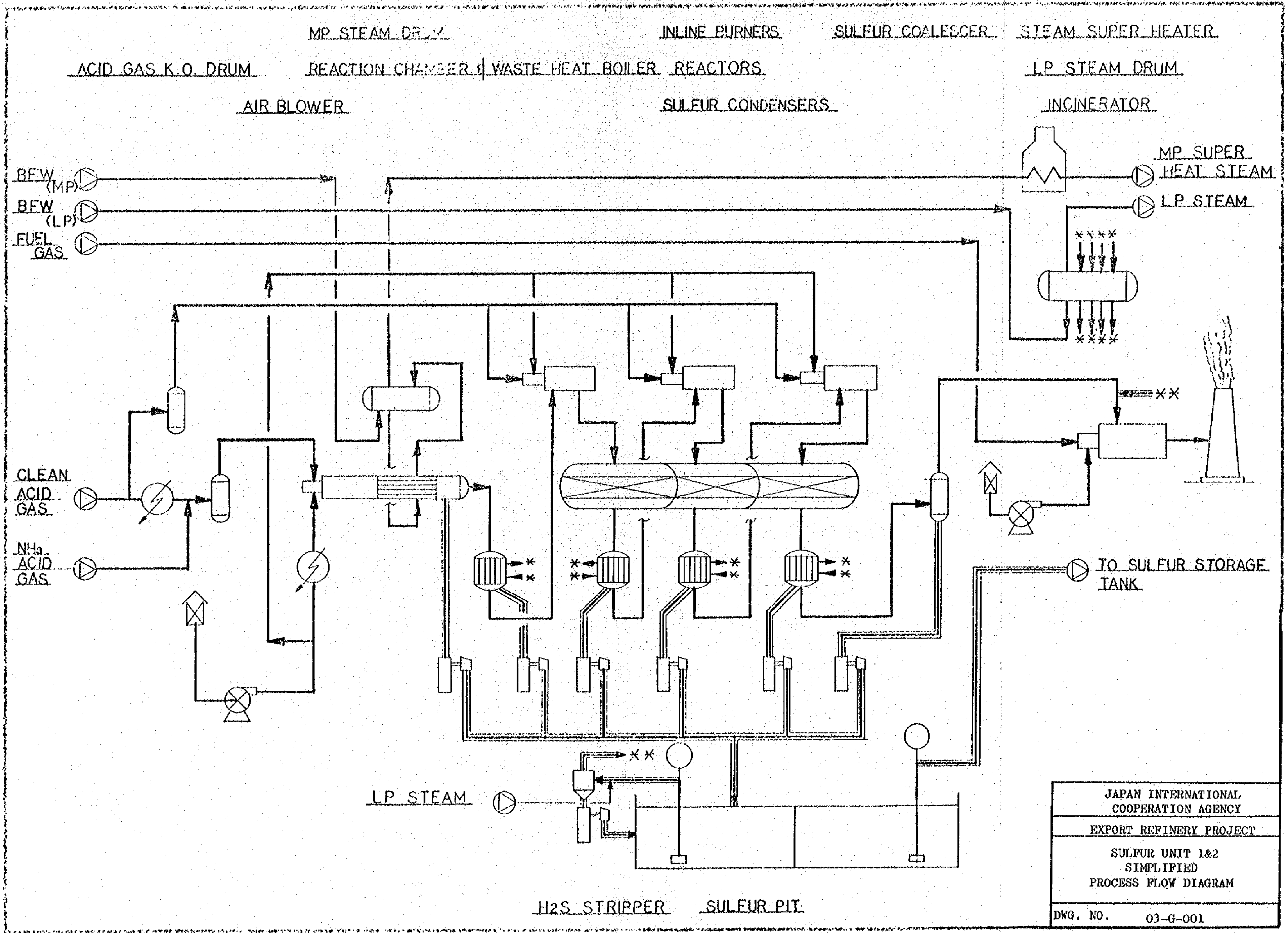
Properties: Purity : 99.9 wt% (Dry) minimum

Color : Bright yellow at ambient temperature

Acidity: Nil

4.12.7 Battery limit conditions

Fluid	Temperature (°C)	Pressure (kg/cm^{2}G)	Note
Acid gas	40	0.6	
NH_3 acid gas	80	0.6	
Sulfur	160	5.0	
Acid gas drain	40	3.0	



JAPAN INTERNATIONAL
COOPERATION AGENCY

EXPORT REFINERY PROJECT

SULFUR UNIT 1&2
SIMPLIFIED
PROCESS FLOW DIAGRAM

DWG. NO. 03-G-001