CHAPTER 4. PROCESS DESCRIPTION

CHAPTER 4. PROCESS DESCRIPTION

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TABLES AND ILLUSTRATIONS

4. Process Description

4-1 Olefin Manufacturing Plant by Steam Cracking

(1) Licensor list

Braun A.G.
Fluor Corp.
Foster Wheeler Corp.
The M. W. Kellogg Co.
Linde A.G.
Lummus Co.
Stone & Webster Engineering Corp.
Universal Oil Products

(2) Process description

Ethylene is so essential as a raw material for the petrochemical industry.

It can be produced through recovery from by-product gas, various cracking processes, coke oven gas or dehydration of ethanol by hydrogenation of acetaldehyde and fermentation. Today, however, most of ethylene in large-scale production is manufactured by thermal cracking of ethane, propane, or butane separated from natural gas, refinery off gas, or light fractions of petroleum like naphtha.

(a) Raw material

In the United States, although natural gas is abundant and by-product gas is available at low cost from domestic refineries, demand for naphtha fraction is great for automobile gasoline use. Accordingly, the overwhelming majority of ethylene plants in this country uses natural gas or refinery off gas as feed.

On the other hand, in Europe and Japan, demands for petroleum product are centered on middle and heavy fractions and, further, gasoline demand is relatively small and natural gas is not abundant. Therefore, most of ethylene plants in Europe and Japan use naphtha as feedstock. Presently, increasing attention is being given to cracking of heavy fractions such as gas oil, vacuum gas oil and crude oil in accordance with the growing shortage of naphtha and natural gas.

When feed is natural gas, as will be explained later, almost nothing but ethylene and propylene is produced from a cracking plant. However, when naphtha is used as feed, butadiene, aromatics, etc., are available as by-products though olefin yield decreases. Therefore, utilization of such by-products could reduce production costs of a naphtha cracking plant, as long as large-scale production is permissible.

(b) Production process

Ethylene manufacturing process using gas as raw materials can be roughly divided into the following three sections:

- Cracking section
- Quenching section
- Purification section

1) Cracking section

The cracking section, cracks feed gas to form olefin-rich gas. Crackers are of several types as given below. For the cracking of gas, naphtha, or gas oil, the external-heat tubular furnace is mostly used and for crude cracking, other types are generally used to avoid coking inside the tubes.

- External-heat tubular furnace
- Heat medium circulation cracker
- Superheated steam (as heat medium) cracker
- Partial combustion furnace
- Flame cracker
- Catalytic cracker

The external-heat tubular furnace is designed to apply heat from outside the tubes in which feedstock flows. The composition of cracked gas depends on cracking temperature, residence time and partial pressures of hydrocarbon (to be reduced with steam). Generally, cracking temperature is as high as 700-900°C and residence time is approximately a fraction of a second to 2 seconds. While cracking takes place inside the tubes, polymerization reactions of cracked products occur forming tar, etc.

2) Quenching section

Cracked gas needs to be quenched to minimize the second order reaction, like polymerization, which forms tar and thereby reduces the yield of olefin including ethylene. Therefore, the quenching section is installed immediately after the cracker.

The quencher is of following types:

- Up flow type
- Down flow type
- Horizontal type
- Melting metal direct flow type

Today, the melting metal direct flow type is rarely used because there is fear that metal fine particles enter cracked gas. In the other three types, quenching is performed when effluent gas runs in the tubes and cooling water runs in the shell and in turn, high-pressure steam is recovered.

3) Recovery and purification section

Quenched cracked gas is sent to a recovery and purification section where C_4 lighter gas is separated from the heavy fuel oil fraction with a gasoline fractionator and from cracked gasoline with a quenching tower.

For the recovery and purification of the separated C_4 lighter gas, the following three means are conceivable:

Low temperature fractionation
Selective absorption
Selective adsorption

Selective absorption and selective adsorption are no longer adopted in newly constructed plants.

Low temperature fractionation is of two types: one under high operating pressures of 30 - 40 kg/cm^2 and the other, under low pressures of 2 - 10 kg/cm^2. Today, large plants generally employ the high pressure process. The C_4 and lighter fractions separated in the quenching tower are further pressurized, washed, dried, cooled, and then separated into hydrogen, methane, ethylene, propylene, and the C_4 fraction with fractionators. During the processing, acetylene in the ethylene and ethane fractions and propadiene in the propylene and propane fractions are hydrogenated. Recovered ethane is recycled to the cracking furnace.

(c) Process flow diagram for ethylene plant (Ref. Fig. 4-1)

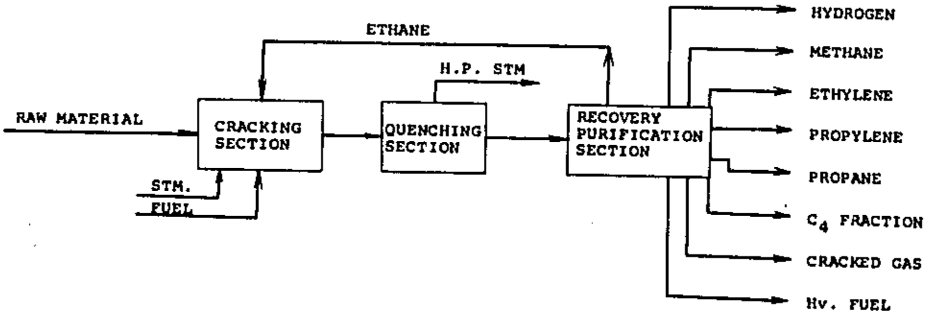


Fig. 4-1 Process Flow Diagram for Ethylene Plant

4-2 Low Density Polyethylene

(1) Licensor list

Imperial Chemical Industries Ltd.
Union Carbide Corp.
E. I. Du Pont De Nemours & Co.
The Dow Chemical Co.
Badische Anilin & Soda-Fabrik A.G.
Dart Industries Inc. (Rexall Drug & Chemical Co.)
U. S. Industries, Inc.
Gulf Oil Corp.
Ethylene Plastics

(2) Process description

In the high pressure polyethylene process, ethylene gas is kept under the conditions of high temperature and pressure and is polymerized in the presence of an initiator. Although polymerization occurs through ordinary radical mechanism, it is characterized with polymerization under very high pressure. Generally, selected pressure and temperature range from 1,000 to 3,000 kg/cm^2 and from 100 to 300°C, respectively.

The polymerization mechanism of ethylene can usually be represented by an elementary equation process in the ordinary radical polymerization.

(a) Decomposition of initiator

$$R_2 \longrightarrow 2R\cdot$$

(b) Initial reaction

$$R + CH_2 = CH_2 \longrightarrow R - CH_2\cdot CH_2$$

(c) Polymerization

$$R - (CH_2\cdot CH_2)_n\ CH_2\cdot CH_2 + CH_2 = CH_2 \longrightarrow R - (CH_2\cdot CH_2)_{n+1}\ CH_2-CH_2$$

(d) Termination of reaction

$$R - (CH_2\cdot CH_2)_n\ CH_2-CH_2\cdot + R - (CH_2\cdot CH_2)_m CH_2-CH_2\cdot \longrightarrow R - (CH_2\cdot CH_2)_{n+1} - (CH_2\cdot CH_2)_{m+1}\ R$$

(e) Transfer reaction

$$R - (CH_2\cdot CH_2)_n\ CH_2-CH_2\cdot + HA \longrightarrow R - (CH_2\cdot CH_2)_n\ CH_2-CH_3 + A\cdot$$

Besides the above reactions, various complicated ones take place.

Since the polymerization of ethylene is such an exothermic reaction as stated below, its yield depends, in fact, on the capability to remove the polymerization heat.

$$CH_2 = CH_2 \longrightarrow \frac{1}{n} (C_2E_4)_n \; - \Delta H = 22.17 \; \frac{Kcal}{mol} \; 500°K$$

The polymerization process comprises the following:

(a) Purification of ethylene gas

(b) Compression and pressurization of ethylene gas

(c) Polymerization

(d) Separation of polymer from unreacted ethylene monomer

(e) Pelletization of polymer

(f) Recycling of ethylene gas

Further details may be summarized in the following paragraphs:

i) Ethylene gas

Ethylene gas to be used as monomer usually requires to have purity of 99.9% or more, and especially content of acetylene should be less than 10 ppm. If mixed in the ethylene gas even in a small amount, CO and aldehyde entail poorer electrical property of the product monomer.

ii) Polymerization

Ethylene gas is pressurized by compressor through a given high polymerization pressure, which, in most cases, ranges from 1,000 to 3,000 kg/cm^{2}g.

There are two types of reactors: one is of a vessel type and the other is of a tubular type. In the case of the former, the pressurized ethylene gas is fed to the reactor after being cooled, while in the latter case, the said gas goes through the preheating system and then is polymerized after being heated up to a given temperature. The polymerization temperature ordinarily ranges from 100 to 300°C. The types and quantity of the initiator to be fed to the reactor vary with polymerization temperature. Peroxide and other compounds which generate free radical are usually used as initiator.

iii) Separation and recycling

Mixture of polyethylene and unreacted ethylene gas which has gone through the reactor is fed to the separator where it is de-pressurized. The reduction in pressure lowers the solubility of ethylene monomer in molten polyethylene with the result of their easy separation. The ethylene separated from polyethylene is recycled to the compressor through a filter.

iv) Pelletization

Polyethylene which has left the separator is fed to the extruder where the said polyethylene is extruded to be pelletized. There are two cutting methods in pelletization: one is "under-water cutting method" which cuts the polymer under water, and the other is "hot cutting method" which cuts it in the atmosphere.

The physical properties and quality of the polyethylene so produced are indicated in terms of molecular weight distribution, density, etc. The density of this polymer, however, is from 0.915 to 0.930, thus, it is lower in density in comparison with the polyethylene manufactured by the medium pressure process described 4-3 hereafter, and is called "low density polyethylene".

(3) Process flow diagram with material balance

Fig. 4-2 shows the process flow diagram with material balance for LDPE Plant.

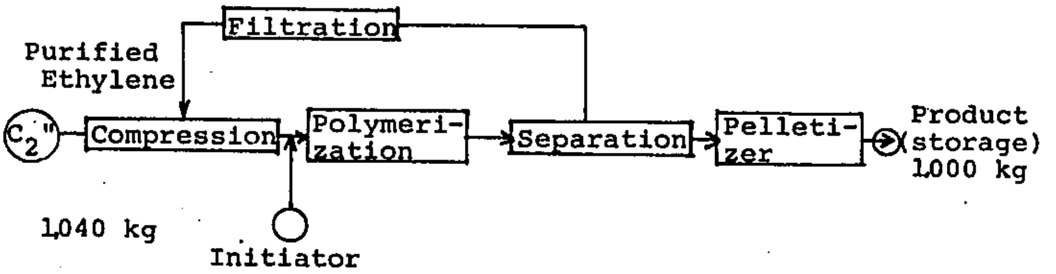


Fig. 4-2 Process Flow with Material Balance for LDPE Plant

4-3 High Density Polyethylene

(1) Licensor list

Mitsui Petrochemical Industries, Ltd.
Union Carbide Corp.
Solvey & Cie, S. A.
Farbwerke Hoechst A.G.
Montedison S.p.A.
Phillips Petroleum Co.
Imperial Chemical Industries Ltd.
SOI
E. I. Du Pont De Nemours & Co.
Hercules Inc.
N. V. Nederlandse Staatsmijnen
Badische Anilin & Soda-Fabrik A.G.

(2) Process description

HDPE processes will be considered by the types of phases present in the polymerization reactor although there were different past classifications. Processes of similar designs can be grouped together as shown in Table 4-1. Now that vapor phase processes are a commercial reality, the list of potential polymerization media is complete. Processes can thus be classified as:

Solution
Slurry
Vapor phase

Table 4-1 Type of HDPE Processes

	Examples
Polymer formed in solution	
Soluble or firmly divided catalyst	Du Pont
Supported catalysts	SOI, Phillips
Polymer forms a slurry with diluent	
Solid catalysts	Ziegler(Mitsui Petrochemical)
Supported catalysts	Solvey, Montedison, Phillips
Polymer forms solid particles suspended in vapor	
Solid catalysts	BASF
Supported catalysts	Union Carbide, BASF

It should be emphasized that the reaction medium used does affect the structure of the polymer, independent of catalyst effects. In turn, this will give rise to basic differences in properties between polymers produced by different processes; these differences can be overcome, in part, by catalyst technology.

In addition, other considerations, such as ease of process control, catalyst efficiency, ability to switch from one grade to another with minimum "twilight" material, process complexity, and product purity must be weighed in the choice of the optimum process.

(a) Solution process (Ref. Fig. 4-3)

In solution processes, ethylene monomer and comonomers are dissolved in hot cyclohexane or other solvent suitable for polyethylene. Catalyst is introduced into the reactor, and the temperature maintained above 140-150°C-the polyethylene melting point at reactor pressure. Some processes use water jacketing to remove reaction heat while others are cooled by monomer refrigeration.

Solution processes are generally run at moderately high pressures and require heavier wall reactors than slurry-type processes. However, because of the beneficial effect of increased temperature on reaction rates, catalyst efficiencies are usually high with short residence times. This allows a higher production rate for a given size reactor. This type of process is inherently limited in the amount of polymer which can be kept in solution--35-40 percent is the maximum. Also, making high molecular weight polymer gives difficulties such as putting high torque on the stirrer, dropping out of solution as a gel, and fouling the reactor. Thus, it is generally more difficult to make extrusion blow molding grades with solution processes, particularly those requiring a very high molecular weight component for high melt strength and die swell.

On the other hand, solution processes generally excel in producing injection molding grades, where narrower molecular weight distribution's and lower molecular weight's are required.

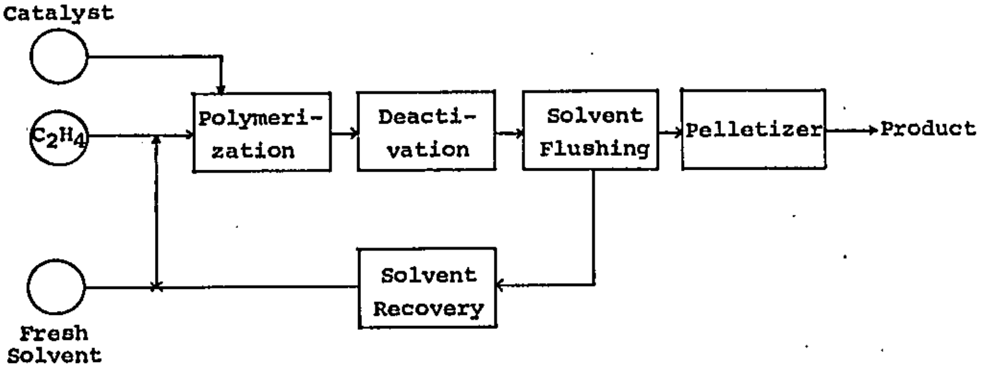


Fig. 4-3 HDPE Solution Process

(b) Slurry (particle form) process (Ref. Fig. 4-4)

In this type of process, a poorer solvent (diluent) for polyethylene is used; the polymer or copolymer is separated from the diluent in fine particles. In this way, the viscosity of the diluent does not increase as rapidly as in the solution processes, and a higher concentration of polymer can be maintained in the reactor. However, many slurry processes are susceptible to fouling--a buildup of polymer adhering to stirrer blades and shafts, reactor walls and flow lines. The time between turnarounds to clean reactors can vary considerably from one process to another, and is, at least in part, dependent on the nature of the catalyst. Some of the advantages of slurry-type processes include a higher volume yield of product for a given reactor size, as well as the greater ease of diluent removal. Another advantage is the potential for making powders suitable for rotamolding directly in the reactor, and thus, cutting out the expensive grinding step.

Reactors used for slurry processes may take any of several forms, from kettles to loop-type designs. The latter has high surface-to-volume ratios, which is necessary for controlling reactor temperature, to keeping molecular weight and molecular weight distribution on target.

In practice, most slurry processes, unless modified, tend to yield very high molecular weight materials which are not commercially useful. Thus, a "chain stopper" or chain transfer agent is often required. For Ziegler-type catalysts, hydrogen is generally used; it tends to give a "clean" product and is not excessively expensive.

Because of their tendency to make higher molecular weights, most slurry processes perform well in making blow molding grades; in some instances, however, tandem reactors may be required for broad molecular weight distribution's.

In general, three types of catalysts are employed in slurry processes: a) pure Ziegler, b) supported Ziegler, and c) Phillips.

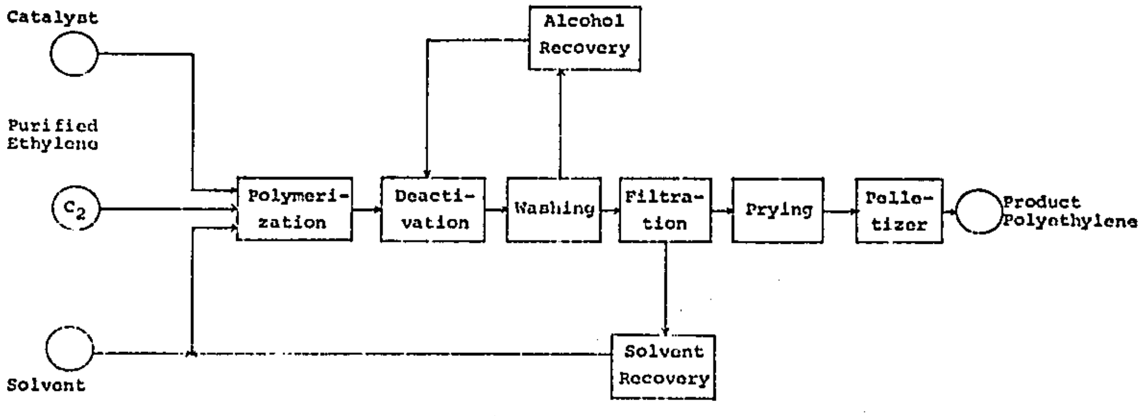


Fig. 4-4 HDPE Slurry Process

(c) Vapor phase process (Ref. Fig. 4-5)

Theoretically, a vapor phase process appears to have the lowest investment and operating costs, provided several conditions could be met. Problems regarding the preparation of a suitable catalyst hindered early commercialization. Excessive catalyst particle size was one difficulty and another was finding catalyst of sufficiently high activities (to eliminate deashing) along with the desired molecular weight and molecular weight distribution. Other problem concerned heat removal and maintenance of constant temperatures.

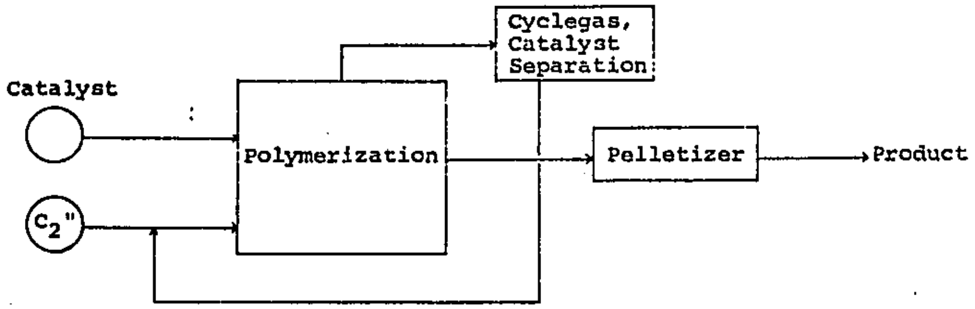


Fig. 4-5 HDPE Vapor Phase Process

4-4 Electrolysis of Salt

(1) Licensor list

(a) Diaphragm cell process

Diamond Shamrock Corp.
Hooker Chemical Corp.
PPG Industries, Inc.
The Dow Chemical Co.
USSR

(b) Mercury cell process

Diamond Shamrock Corp.
Hooker Chemical Corp.
Asahi Glass Co., Ltd.
De Nora Oronzio S.p.A.
Kureha Chemical Industry Co., Ltd.
Krebs-BASF
Krebs-Zuerich
Mitsui Toatsu Chemicals, Inc.
Olin Corp.
PPG Industries, Inc.
Solvay & Cie, S. A.
Toyo Soda Manufacturing Co., Ltd.
Uhde-Hoechst

(c) Ion exchange membrane cell process

Asahi Chemical Industry Co., Ltd.

(2) Process description

Salt (sodium chloride) in an aqueous solution dissociated to give sodium and chloride ions. When cathode and anode are placed in the aqueous solution and then direct current is supplied between the electrodes, the chlorine ion moves toward the anode and the sodium ion toward the cathode. At the cathode, the sodium ion discharges and at the same time reacts with water to give sodium hydroxide (caustic soda) and hydrogen gas. The chloride ion is converted into chlorine gas by discharge at the anode.

$$Na^+ + H_2O + e^- \longrightarrow NaOH + \tfrac{1}{2}H_2$$

$$2\,Cl^- \longrightarrow Cl_2 + e^-$$

When mixing takes place between the cathode and anode, different reactions occur and also the efficiency of the current decreases. In order to avoid this undesirable process, three methods are adopted in the industry. They are 'Mercury process', 'Diaphragm process', and 'Ion exchange membrane process'.

In the diaphragm process, the mixing of the polar components is inhibited by inserting a diaphragm made of asbestos between the cathode and anode. The concentration of sodium hydroxide in the electrolyte is low (12%), and also the solution contains about 17% of sodium chloride. Moreover, caustic soda concentration and salt separation process by vacuum evapolation is necessary.

In the mercury process, mercury cathode is used. At the cathode, discharged sodium reacts with mercury to form sodium amalgam, which is transferred into another vessel. Decomposition of the amalgam with water gives sodium hydroxide and hydrogen.

$$Hg\text{-}Na + H_2O \longrightarrow Hg + NaOH + \tfrac{1}{2}\,H_2$$

High concentration (50%) of sodium hydroxide is obtained by the mercury process. Since sodium amalgam is separated from salt solution and decomposed with water, low chloride content caustic soda is produced. On the other hand, caustic soda and waste materials are contaminated with mercury.

Generally speaking, the mercury process is superior economically to the diaphragm process and industrial production of sodium hydroxide is carried out mostly by the mercury process. However, pollution by mercury is now a serious ecological problem, and at present the diaphragm process is replacing the mercury process.

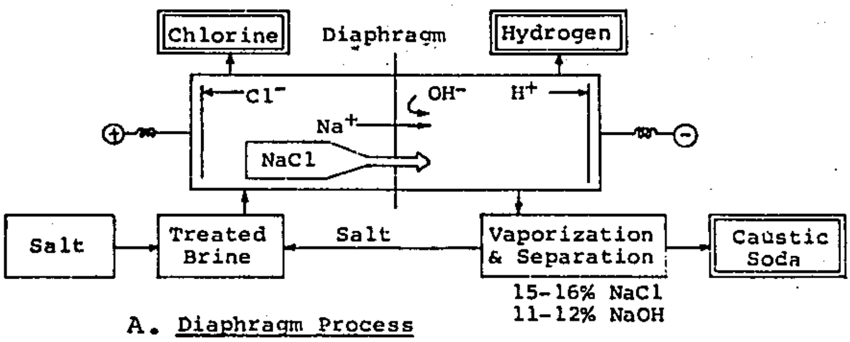
However, to avoid the economic disadvantage of caustic soda rectifying in the diaphragm cell process, much effort is being made in the development of an ion exchange membrane cell process.

The contamination of caustic soda with salt is prevented by inserting ion exchange membrane,of mainly floric plastics, between cathode and anode, so that the property of caustic soda is nearly equivalent to that of caustic soda produced by the mercury process.

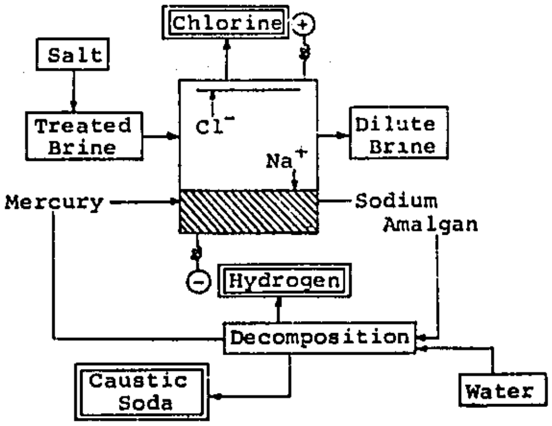
It is reported that ion exchange membrane cell process has:

(a) the advantage to produce high purity caustic soda and

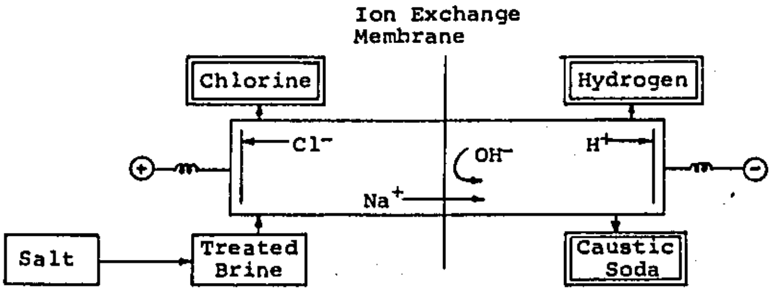
(b) a maximum production capacity of a single train of electrolytic cell. Thus, production unit is compact and construction is at low cost.



A. Diaphragm Process



B. Mercury Process



C. Ion Exchange Membrane Process

Fig. 4-6 Diagrams of Electrolysis Processes

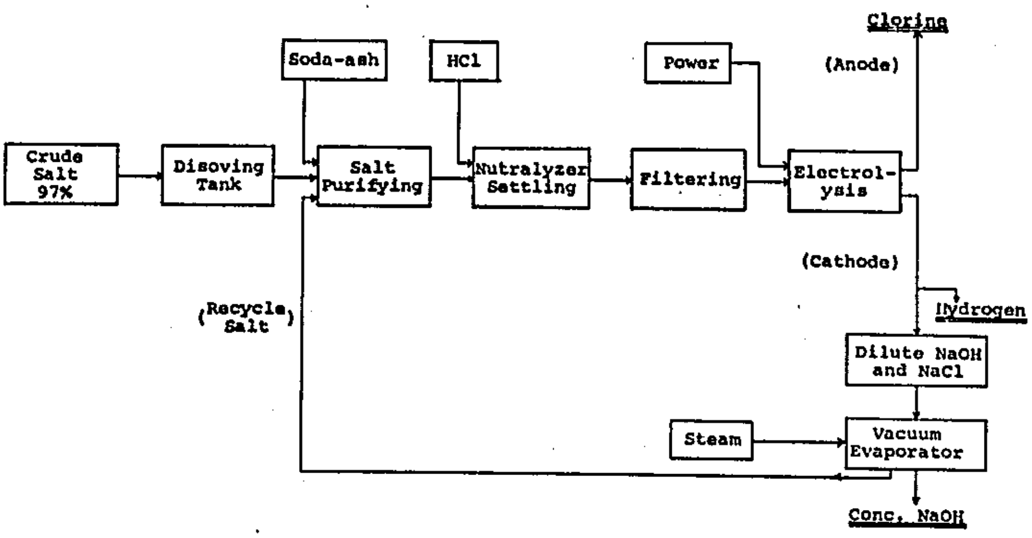


Fig. 4-7 Process Flow Sheet of Electrolysis of Salt by Diaphragm Cell Process

4-5 Vinyl Chloride Monomer

(1) Licensor list

Chemische Werke Huels A.G.
Farbwerke Hoechst A.G.
Kureha-Chiyoda
Mitsui Toatsu Chemicals, Inc.
Monsanto Co.
Solvey ICI
Toyo Soda Manufacturing Co., Ltd.
Tokuyama Soda Co., Ltd.
Union Carbide Corp.
The B. F. Goodrich Co.
Montedison S.p.A.
Scientific Design Co., Inc.
Stauffer Chemical Co.
PPG Industries, Inc.

(2) Process description

The VCM plant consists mainly of the following sections:

Direct chlorination (EDC)
Oxychlorination (EDC)
EDC purification
EDC cracking —— VCM purification —— Product VCM

(a) Direct chlorination

Ethylene and chlorine are fed into the reactor where the reaction takes plance in the liquid phase of EDC at its boiling point.

$$C_2H_4 + Cl_2 \longrightarrow C_2H_4Cl_2$$

The heat of reaction is removed by vaporization of EDC.

(b) Oxychlorination

Oxygen, ethylene and hydrogen chloride are mixed with the recycled gas (mentioned below) and the mixed gases are fed into the reactor with a newly developed fluicized bed.

The composition of raw material gases is controlled to be kept in the nonflammable region. The reaction is carried out at a moderate pressure.

$$C_2H_4 + 2HCl + \frac{1}{2}O_2 \longrightarrow C_2H_4Cl_2 + H_2O$$

The heat of reaction is removed effectively by circulating pressure water which is recovered as steam. The reaction mixture is cooled down, and both EDC and water produced are condensed in a quencher. Non-condensed gases are recycled back to the reactor. EDC is easily separated from the water by decantation, and purified after caustic treatment.

(c) EDC purification

EDC from oxychlorination is dehydrated, and then purified together with EDC from direct chlorination through distillation.

(d) EDC cracking

The purified EDC is cracked to VCM and hydrogen chloride under high temperature and high pressure condition.

$$C_2H_4Cl_2 \longrightarrow C_2H_3Cl + HCl$$

The hot effluents from the cracker are quenched in the EDC quencher.

(e) VCM purification

At first, the hydrogen chloride is separated from effluents in HCl column, and returned to oxychlorination section. Then, EDC is separated from VCM in VCM column and recycled to the EDC purification section. VCM is dried through caustic column. VCM product is delivered to the polymerization section.

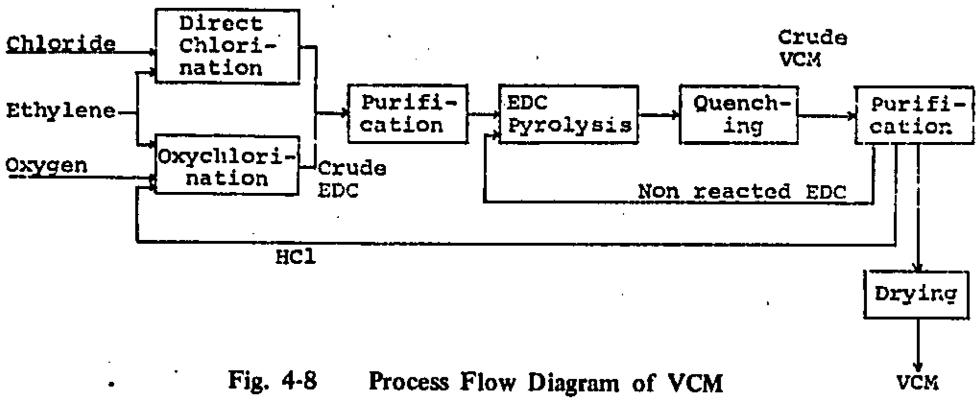


Fig. 4-8 Process Flow Diagram of VCM

4-6 Polypropylene

(1) Licensor list

Montedison S.p.A.
Mitsui Petrochemical Industries, Ltd.
Phillips Petroleum Co.
Badische Anilin & Soda-Fabrik A.G.
SOI
Imperial Chemical Industries Ltd.
Hercules Inc.
Friedrich Uhde GmbH
Farbwerke Hoechst A.G.
Hüls/Veba Chemie
Mitsui Toatsu Chemicals, Inc.

(2) Process description

(a) General

Propylene polymerizes with the Ziegler Natta catalyst system and produces polypropylene. Propylene used for this feed must be free from acetylenic compounds and propadiene. Thus, propylene is rectified and hydrogenated before feeding. (As for the purity of the feed, there are two processes here: one uses extremely high purity propylene with solvents (hexane or heptane) while another uses propylene wherein only acetylenics and propadiene are removed but propane exists without solvents. In the latter case, a big propylene rectifier is not necessary but the propylene is wasted with the purged propane.)

Polypropylene has a branch of the methyl group, and thus, forms some part of the atactic arrangement of molecules. It is not valuable itself as a plastic and excess atactic polymer is separated by the solubility of the solvent.

There are co-polymers of propylene and ethylene other than propylene-homopolymer in the category of polypropylene. They are an improvement on polymer characteristics for tenacity of strength at low temperature services.

Within the past two years many major advances in catalyst and polymerization technology have been made to diminish the atactic formation and deashing the catalyst. One of these developments is the discovery of some high active catalysts.

(b) Process description

Typical example of a process with high purity propylene and solvent is illustrated on the diagram given below.

Catalysts are prepared in the catalyst preparation drum, and propylene is fed into the reactor system with solvent. In case of co-polymer, the ethylene is fed into a series of reactors.

Formed polypropylene slurry is put into the deactivator where the catalysts are deactivated with alcohol and then washed with water.

Alcohol goes into the water stream and it is distilled to recover alcohol. Hexane-polymer slurry is separated from the water-alcohol solution by gravity. Then, polymer is separated from hexane by the decantor. The mother liquor, which contains hexane and atactic polymer, is put into the evaporator where solvent is removed from atactics.

Isotactic polymer, separated by the decantor, is dried in the dryer and becomes polypropylene powder. This powder is pelletized with extruders.

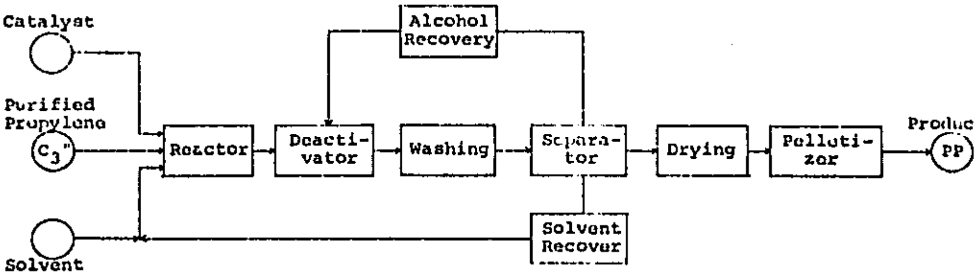


Fig. 4-9 Process Flow Diagram for Polypropylene Plant

CHAPTER 5. FINANCIAL ANALYSIS

CHAPTER 5. FINANCIAL ANALYSIS

CONTENTS

TABLES AND ILLUSTRATIONS

5. Financial Analysis

5-1 Basic Assumptions for Financial Analysis

(1) Investment schedule

Disbursement schedule on construction cost during the pre-operating period is based on the following hypothetical conditions:

	Construction Cost (Progress Amount)	Downpayment	Total Disbursement on Construction Cost
1976	5 %	15 %	19.25 %
1977	20 %	-	17.00 %
1978	40 %	-	34.00 %
1979	35 %	-	29.75 %
Total	100 %		100.00 %

The total construction period of all the plants is assumed to be 3.5 years. However, in the case of the utility center, this is assumed to be only 2.5 years due to a shorter time delivery of equipment and an easier construction job.

(2) Guarantee fee

A guarantee fee on a long-term loan granted by a foreign financier is calculated in accordance with the DBP's "Manual of Operation of Industrial Projects Department I", in which the important provisions are as follows:

(a) Commitment fee : Loan amount x 1.0 %

(b) Guarantee fee on availed amount: Availed amount x 1.5 %

(c) Guarantee fee on unavailed amount : Unavailed amount x 1.0 %

Basically, the guarantee fee during the operating years is calculated according to the above provision (b) but the availed amount shall include payable interest on a long-term loan for each year.

(3) Production and sales schedule

Annual production and sales amount in domestic and export markets are as per attached: Tables 5-2 ~ 5-8 hereof.

(4) Working capital

(a) Working capital is not included in total capital investment figure. Instead, it is financed through a local loan in which repayment is due after 365 days with an 18 percent interest rate.

(b) Except for the HDPE plant, the amount of initial working capital is calculated as follows:

(Accounts receivable in 1982 - Accounts payable in 1982) x 2/12

On the other hand, the working capital of the HDPE plant depends on the respective requirements of cash outlay as shown in the Cash Flow Statement in Table 5-11 of Part I.

(c) The amount of working capital financed in the operating period is decreased according to the following schedule:

1979	100 %
1980	100 %
1981	100 %
1982	100 % x 2/3
1983	100 % x 1/3
1984~	0

(5) Privilege on income tax

In accordance with the provisions of the "Investments Incentives Act", the following items are deducted from taxable income.

(a) Pre-operating expenses

a. Organizational expenses

b. Start-up expenses

(b) Direct labor cost (for export)

(c) Net operating loss

(6) Production cost calculation

(a) Evaluation of by-products

1) Ethylene plant

a) Propylene for LPG

Propylene for LPG is credited back to the Filoil Refinery at a cost of US$122.20/t.

b) LPG

Same as a), i.e., LPG is credited back to the refinery at the amount of US$122.20/t.

c) Other by-products

Other by-products are mostly fuels, which are supplied to downstream plants as well as to the utility center.

2) Electrolysis plant

a) Caustic soda

Caustic soda from the electrolysis plant is sold to the domestic market at a price of US$417/t.

b) Hydrogen

Hydrogen from the electrolysis plant is sold to the refinery at a price of US$261/t.

(7) Delivery condition of products

(a) Polymers

All polymer products from the downstream plants are sold in paper bag containers on an ex-factory warehouse basis.

(b) VCM and caustic soda

VCM and caustic soda are sold in liquid forms on an ex-factory storage basis.

(8) Depreciation (Straight)

(a) All plants except the utility center

a. Process facility : 10 years

b. Off-site facility : 15 years

(b) Utility center

Whole facility : 15 years

(9) Amortization

The following items are straightly amortized for a period of 10 years.

(a) Pre-operating expenses

a. Start-up expenses

b. Organizational expenses

(b) Financial charges during pre-operating period

a. Guarantee fee

b. Interest during construction on long-term loan

(c) License and know-how fee

(10) Maintenance and repair cost

Maintenance and repair cost is calculated as:

[(Process facility x 0.03) + (Off-site facility x 0.01)] x 0.01

- However, the maintenance and repair cost of the first two years are one-third and two-thirds of the total amount, respectively.

(11) Inventory

(a) Raw materials

Only raw salt, in the amount which used during one month's consumption in the electrolysis plant, is taken into consideration.

(b) Products

No inventory is taken to account for any products.

(12) Sales tax

(a) All products sold from the complex to the market are subject to the imposition of sales tax. Computations are based on the following rates considered to be one of the privileges given to pioneer industries:

1980	0.875	1986	5.950
1981	1.750	1987	6.300
1982	1.750	1988	6.300
1983	2.625	1989	6.650
1984	3.500	1990	7.000
1985	4.550		

(b) Any sales and/or procurements among the plants within the complex, including the refinery, are assumed to be free of sales tax or specific tax charges.

(13) Income tax

A thirty-five percent (35%) income tax rate is imposed on taxable income.

(14) Accounts receivable

For the sale of products, the payment due date is assumed to be sixty (60) days after delivery. Thus, accounts receivable is computed as:

Total sales x 2/12

(15) Accounts payable

In the procurement of raw materials, utilities, etc., the same payment period for the accounts receivable is followed. Therefore, accounts payable is calculated as:

Total variable cost x 2/12

(16) Administrative expenses and contingency

Administrative and contingency expences for the operating years are calculated as follows:

(a) Administrative expenses

(Total production cost - Contingency) ---------------a

(Total operating expenses - Administrative expenses)-b

(a + b) x 0.03

(b) Contingency

(a + b + Administrative expenses) x 0.02

(17) Real property tax and insurance

(a) Real property tax

(Total fixed assets - Land) x 0.008 + (Land x 0.01)

(b) Insurance

(Total fixed assets - Land) x 0.006

(18) Internal rate of return on discounted cash flow method (IRR)

In accordance with the BOI's practice, the IRR is calculated based on the following formula.

(a) IRR on total capital requirement

$$C_t = \sum_{n=0}^{10} \frac{C_n}{(1 + r)^n}$$

where

C_t: Total capital requirement or total project cost

C_n: Cash inflow including net cash flow, long-term loan repayment, interest on long-term loan, other financial charges for long-term loan, and working capital

(b) IRR on paid-in capital

$$E = \sum_{n=0}^{10} \frac{NCI}{(1 + r)^n}$$

where

E : Paid-in capital

NCI : Net cash inflow

5-2 Additional Information

5-2-1 Capital Requirement

The capital requirement for the project is as shown in the Table 5-1.

Table 5-1 Summary of Capital Requirement

(1,000 US$)

Item / Name of Plant (Rated Capacity)	Ethylene Plant (303,085 MT/y)	Utility Center (various)	LDPE Plant (110,000 MT/y)	HDPE Plant (60,000 MT/y)	Electrolysis Plant (Cl_2 37,540 MT/y)	V C M Plant (55,000 MT/y)	PP Plant (90,000 MT/y)	Total
Construction Cost								
ISBL Facility	154,250	71,770	79,320	33,480	22,860	21,240	60,560	443,480
Off-site Facility	13,410	12,440	9,120	5,895	2,415	2,760	8,475	54,515
sub-total	167,660	84,210	88,440	39,375	25,275	24,000	69,035	497,995
Other Supplies								
Spare Parts	5,400	4,440	2,780	1,170	800	930	2,120	17,640
Catalyst and Chemicals	680	-	-	-	-	300	-	980
Licence and Knowhow Fee	1,920	-	5,960	5,060	1,810	700	7,640	23,090
sub-total	8,000	4,440	8,740	6,230	2,610	1,930	9,760	41,710
Pre-operational Expense								
Start-up Expense	6,030	1,530	3,630	2,230	800	1,340	2,890	18,450
Organizational Expense	4,470	2,120	2,500	1,200	740	680	2,020	13,730
sub-total	10,500	3,650	6,130	3,430	1,540	2,020	4,910	32,180
Financial Charge								
Guarantee Fee	8,740	3,260	4,090	2,380	1,450	1,330	3,980	26,030
Interest during Construction	25,740	10,000	14,620	7,270	4,430	3,890	12,090	78,040
sub-total	34,480	13,260	19,510	9,650	5,880	5,220	16,070	104,070
Land	2,954	500	1,985	1,560	1,560	945	1,415	10,919
Total Investment Cost	223,594	106,060	124,805	60,245	36,865	34,115	101,190	686,874
Working Capital	12,551	4,873	6,832	4,000	2,065	2,124	7,694	40,139
Total Capital Requirement	236,145	110,933	131,637	64,245	38,930	36,239	108,884	727,013
Finance								
Equity Capital	67,074	31,820	37,445	18,075	11,065	10,235	30,350	206,064
Foreign Loan	156,520	74,240	67,360	42,170	25,800	23,880	70,840	480,810
sub-total	223,594	106,060	124,805	60,245	36,865	34,115	101,190	686,874
Local Loan	12,551	4,873	6,832	4,000	2,065	2,124	7,694	40,139
Total Finance	236,145	110,933	131,637	64,245	38,930	36,239	108,884	727,013

5-2-2 Production and Sales Schedule

The projected production and sales schedule is as shown in Tables 5-2 through 5-8:

Table 5-2	Ethylene plant
Table 5-3	Utility center
Table 5-4	LDPE plant
Table 5-5	HDPE plant
Table 5-6	Electrolysis plant
Table 5-7	VCM plant
Table 5-8	PP plant

Table 5-2 Production and Sales Projection
Ethylene Plant

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
Production										
C_2 (MT)	121,859	152,323	182,788	182,788	182,788	182,788	182,788	182,788	182,788	182,788
C_3 (MT)	61,561	76,949	92,340	92,340	92,340	92,340	92,340	92,340	92,340	92,340
Fuel (MT)	248,087	310,108	372,130	372,130	372,130	372,130	372,130	372,130	372,130	372,130
Sales										
Domestic										
C_2 (MT) @ US$557.7/MT	121,859	152,323	182,788	182,788	182,788	182,788	182,788	182,788	182,788	182,788
Sales (1,000 US$)	67,961	84,951	101,941	101,941	101,941	101,941	101,941	101,941	101,941	101,941
C_3 (MT) @ US$446.19/MT	61,561	76,949	92,340	92,340	92,340	92,340	92,340	92,340	92,340	92,340
Sales	27,468	34,334	41,201	41,201	41,201	41,201	41,201	41,201	41,201	41,201
Fuel (MT) @ US$138.5/MT	248,087	310,108	372,130	372,130	372,130	372,130	372,130	372,130	372,130	372,130
Sales	34,360	42,950	51,540	51,540	51,540	51,540	51,540	51,540	51,540	51,540
Export	0	0	0	0	0	0	0	0	0	0
Total Sales	129,789	162,235	194,682	194,682	194,682	194,682	194,682	194,682	194,682	194,682

Table 5-3 Production and Sales Projection
Utility Center

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
<u>Production and Sales</u>										
High Pressure Steam (MT)	4	5	6	6	6	6	6	6	6	6
@ US$15.1/MT	483	604	725	725	725	725	725	725	725	725
Middle Pressure Steam (MT)	116	146	175	175	175	175	175	175	175	175
@ US$11.6/MT	10,765	13,549	16,240	16,240	16,240	16,240	16,240	16,240	16,240	16,240
Low Pressure Steam (MT)	-6	-8	-9	-9	-9	-9	-9	-9	-9	-9
@ US$6.52/MT	-313	-417	-469	-469	-469	-469	-469	-469	-469	-469
Electric Power (KWH)	35,725	44,657	53,588	53,588	53,588	53,588	53,588	53,588	53,588	53,588
@ US$0.067/KWH	19,149	23,936	28,723	28,723	28,723	28,723	28,723	28,723	28,723	28,723
Well Water (MT)	539	674	809	809	809	809	809	809	809	809
@ US$0.0543/MT	234	293	351	351	351	351	351	351	351	351
Sea Water (MT)	6,795	8,494	10,193	10,193	10,193	10,193	10,193	10,193	10,193	10,193
@ US$0.103/MT	5,599	6,999	8,399	8,399	8,399	8,399	8,399	8,399	8,399	8,399
Boiler Feed Water (MT)	5	6	7	7	7	7	7	7	7	7
@ US$1.01/MT	40	48	57	57	57	57	57	57	57	57
Instrument Air (MT)	2,899	3,624	4,349	4,349	4,349	4,349	4,349	4,349	4,349	4,349
@ US$0.0257/MT	596	745	894	894	894	894	894	894	894	894
Inert Gas (MT)	2,471	3,089	3,707	3,707	3,707	3,707	3,707	3,707	3,707	3,707
@ US$0.127/MT	2,511	3,138	3,766	3,766	3,766	3,766	3,766	3,766	3,766	3,766
Plant Air (m^3)	1,396	1,745	2,093	2,093	2,093	2,093	2,093	2,093	2,093	2,093
@ US$0.022/$m^3$	246	307	368	368	368	368	368	368	368	368
<u>Total Sales</u>	39,310	49,202	59,054	59,054	59,054	59,054	59,054	59,054	59,054	59,054

Table 5-4 Production and Sales Projection
LDPE Plant

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
<u>Production</u>										
LDPE (MT)	66,000	82,500	99,000	99,000	99,000	99,000	99,000	99,000	99,000	99,000
<u>Sales</u>										
Domestic										
Quantity (MT) @ US$1,347.7/MT	47,367	54,230	62,088	71,084	81,385	93,177	99,000	99,000	99,000	99,000
Sales (1,000 US$)	63,835	73,085	83,675	95,800	109,681	125,574	133,421	133,421	133,421	133,421
Export										
Quantity (MT) @ US$681.0/MT	18,634	28,270	36,912	27,916	17,615	5,823	0	0	0	0
Sales (1,000 US$)	12,689	19,252	25,137	19,011	11,996	3,965	0	0	0	0
<u>Total Sales</u>	76,525	92,337	108,812	114,810	121,677	129,539	133,421	133,421	133,421	133,421

Table 5-5 Production and Sales Projection
HDPE Plant

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
Production										
HDPE (MT)	36,000	45,000	54,000	54,000	54,000	54,000	54,000	54,000	54,000	54,000
Sales										
Domestic										
Quantity (MT) @US$1347.7/MT	23,449	26,847	30,737	35,190	40,289	46,127	52,811	54,000	54,000	54,000
Sales(1,000 US$)	31,602	36,181	41,423	47,426	54,298	62,165	71,173	72,775	72,775	72,775
Export										
Quantity (MT) @ US$773.0/MT	12,551	18,153	23,263	18,810	13,711	7,873	1,189	0	0	0
Sales(1,000 US$)	9,702	14,033	17,983	14,540	10,598	6,086	919	0	0	0
Total Sales	41,304	50,213	59,406	61,965	64,896	68,251	72,092	72,775	72,775	72,775

Table 5-6 Production and Sales Projection
Electrolysis Plant

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
Production										
Cl_2 (MT)	19,536	24,420	29,304	29,304	29,304	29,304	29,304	29,304	29,304	29,304
NaOH (MT)	22,175	27,717	33,264	33,264	33,264	33,264	33,264	33,264	33,264	33,264
Sales										
Domestic										
Cl_2 (MT) @ US$255.7/MT	19,536	24,420	29,304	29,304	29,304	29,304	29,304	29,304	29,304	29,304
Sales (1,000 US$)	4,995	6,244	7,493	7,493	7,493	7,493	7,493	7,493	7,493	7,493
NaOH (MT) @ US$417/MT	22,175	27,717	33,264	33,264	33,264	33,264	33,264	33,264	33,264	33,264
Sales (1,000 US$)	9,247	11,558	13,871	13,871	13,871	13,871	13,871	13,871	13,871	13,871
Export	0	0	0	0	0	0	0	0	0	0
Total Sales	14,242	17,802	21,364	21,364	21,364	21,364	21,364	21,364	21,364	21,364

Table 5-7 Production and Sales Projection
VCM Plant

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
Production										
VCM (MT)	33,000	41,250	49,500	49,500	49,500	49,500	49,500	49,500	49,500	49,500
Sales										
Domestic										
VCM (MT)	33,000	41,250	49,500	49,500	49,500	49,500	49,500	49,500	49,500	49,500
@US$691.7										
Sales(1,000 US$)	22,826	28,533	34,239	34,239	34,239	34,239	34,239	34,239	34,239	34,239
Export	0	0	0	0	0	0	0	0	0	0
Total Sales	22,826	28,533	34,239	34,239	34,239	34,239	34,239	34,239	34,239	34,239

Table 5-8 Production and Sales Projection
PP Plant

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
Production										
PP (MT)	54,000	67,500	81,000	81,000	81,000	81,000	81,000	81,000	81,000	81,000
Sales										
Domestic										
PP (MT)	45,708	51,908	58,949	66,945	76,026	81,000	81,000	81,000	81,000	81,000
@ US$1371.5/MT										
Sales(1,000 US$)	62,691	71,194	80,851	91,818	104,273	111,095	111,095	111,095	111,095	111,095
Export										
PP (MT)	8,292	15,592	22,051	14,055	4,974	0	0	0	0	0
@ US$813.0/MT										
Sales(1,000 US$)	6,741	12,676	17,927	11,426	4,044	0	0	0	0	0
Total Sales	69,432	83,870	98,779	103,245	108,317	111,095	111,095	111,095	111,095	111,095

5-2-3 Projected Cost of Goods Manufactured and Sold

The projected cost of goods manufactured and sale is as shown in Tables 5-9 through 5-15:

Table 5-9 Ethylene plant
Table 5-10 Utility center
Table 5-11 LDPE plant
Table 5-12 HDPE plant
Table 5-13 Electrolysis plant
Table 5-14 VCM plant
Table 5-15 PP plant

Table 5-9 Cost of Goods Manufactured and Sold

Ethylene Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
VARIABLE COST										
Raw Materials	65,896	82,369	98,843	98,843	98,843	98,843	98,843	98,843	98,843	98,843
By-products (credit)	(34,360)	(42,950)	(51,540)	(51,540)	(51,540)	(51,540)	(51,540)	(51,540)	(51,540)	(51,540)
Utilities	17,871	22,338	26,806	26,806	26,806	26,806	26,806	26,806	26,806	26,806
Sub-total	49,407	61,757	74,109	74,109	74,109	74,109	74,109	74,109	74,109	74,109
FIXED COST										
Direct Labor	647	647	647	647	647	647	647	647	647	647
Depreciation	16,927	16,927	16,927	16,927	16,927	16,927	16,927	16,927	16,927	16,927
Repair & Maintenance	1,512	3,023	4,535	4,535	4,535	4,535	4,535	4,535	4,535	4,535
Real Property Tax & Insurance	2,435	2,198	1,961	1,724	1,487	1,250	1,013	776	539	303
Plant Overhead	129	129	129	129	129	129	129	129	129	129
Sub-total	21,650	22,924	24,199	23,962	23,725	23,488	23,251	23,014	22,777	22,541
CONTINGENCY	1,514	1,787	2,060	2,055	2,050	2,046	2,041	2,036	2,032	2,027
TOTAL PRODUCTION COST	72,571	86,468	100,368	100,126	99,884	99,643	99,401	99,159	98,918	98,677
PRODUCTION AMOUNT (MT)	-	-	-	-	-	-	-	-	-	-
BEGINNING INVENTORY	-	-	-	-	-	-	-	-	-	-
ENDING INVENTORY	-	-	-	-	-	-	-	-	-	-
COST OF GOODS SOLD	72,571	86,468	100,368	100,126	99,884	99,643	99,401	99,159	98,918	98,677

Table 5-10 Cost of Goods Manufactured and Sold
Utility Center

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
VARIABLE COST										
Raw Materials	-	-	-	-	-	-	-	-	-	-
By-products (credit)	-	-	-	-	-	-	-	-	-	-
Utilities	21,583	26,760	32,253	32,253	32,253	32,253	32,253	32,253	32,253	32,253
Sub-total	21,583	26,760	32,253	32,253	32,253	32,253	32,253	32,253	32,253	32,253
FIXED COST										
Direct Labor	624	624	624	624	624	624	624	624	624	624
Depreciation	5,910	5,910	5,910	5,910	5,910	5,910	5,910	5,910	5,910	5,910
Repair & Maintenance	759	1,518	2,277	2,277	2,277	2,277	2,277	2,277	2,277	2,277
Real Property Tax & Insurance	1,242	1,159	1,076	993	910	828	745	662	580	497
Plant Overhead	125	125	125	125	125	125	125	125	125	125
Sub-total	8,660	9,336	10,012	9,929	9,846	9,764	9,681	9,598	9,516	9,433
CONTINGENCY	662	783	910	908	906	905	903	901	900	898
TOTAL PRODUCTION COST	30,905	36,879	43,175	43,090	43,005	42,922	42,837	42,752	42,669	42,584
PRODUCTION AMOUNT (MT)	-	-	-	-	-	-	-	-	-	-
BEGINNING INVENTORY	-	-	-	-	-	-	-	-	-	-
ENDING INVENTORY	-	-	-	-	-	-	-	-	-	-
COST OF GOODS SOLD	30,905	36,879	43,175	43,090	43,005	42,922	42,837	42,752	42,669	42,584

Table 5-11 Cost of Goods Manufactured and Sold
LDPE Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
VARIABLE COST										
Raw Materials	39,332	49,165	58,998	58,998	58,998	58,998	58,998	58,998	58,998	58,998
By-products (credit)	-	-	-	-	-	-	-	-	-	-
Utilities	8,158	8,158	8,158	8,158	8,158	8,158	8,158	8,158	8,158	8,158
Sub-total	47,490	57,323	67,156	67,156	67,156	67,156	67,156	67,156	67,156	67,156
FIXED COST										
Direct Labor	612	612	612	612	612	612	612	612	612	612
Depreciation	8,818	8,818	8,818	8,818	8,818	8,818	8,818	8,818	8,818	8,818
Repair & Maintenance	784	1,569	2,353	2,353	2,353	2,353	2,353	2,353	2,353	2,353
Real Property Tax & Insurance	1,279	1,156	1,032	909	785	662	534	415	292	168
Plant Overhead	122	122	122	122	122	122	122	122	122	122
Sub-total	11,615	12,277	12,937	12,814	12,690	12,567	12,439	12,320	12,197	12,073
CONTINGENCY	1,317	1,557	1,785	1,805	1,828	1,859	1,899	1,906	1,904	1,911
TOTAL PRODUCTION COST	60,422	71,112	81,878	81,775	81,674	81,582	81,494	81,382	81,257	81,140
PRODUCTION AMOUNT (MT)	-	-	-	-	-	-	-	-	-	-
BEGINNING INVENTORY	-	-	-	-	-	-	-	-	-	-
ENDING INVENTORY	-	-	-	-	-	-	-	-	-	-
COST OF GOODS SOLD	60,422	71,112	81,878	81,775	81,674	81,582	81,494	81,382	81,257	81,140

Table 5-12 Cost of Goods Manufactured and Sold
HDPE Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
VARIABLE COST										
Raw Materials	22,951	28,688	34,426	34,426	34,426	34,426	34,426	34,426	34,426	34,426
By-products (credit)	-	-	-	-	-	-	-	-	-	-
Utilities	5,319	6,649	7,979	7,979	7,979	7,979	7,979	7,979	7,979	7,979
Sub-total	28,270	35,337	42,405	42,405	42,405	42,405	42,405	42,405	42,405	42,405
FIXED COST										
Direct Labor	424	424	424	424	424	424	424	424	424	424
Depreciation	3,858	3,858	3,858	3,858	3,858	3,858	3,858	3,858	3,858	3,858
Repair & Maintenance	338	675	1,013	1,013	1,013	1,013	1,013	1,013	1,013	1,013
Real Property Tax & Insurance	570	516	462	408	354	300	246	192	138	84
Plant Overhead	85	85	85	85	85	85	85	85	85	85
Sub-total	5,275	5,558	5,842	5,788	5,734	5,680	5,626	5,572	5,518	5,464
CONTINGENCY	753	915	1,069	1,080	1,093	1,109	1,132	1,137	1,136	1,140
TOTAL PRODUCTION COST	34,298	41,810	49,316	49,273	49,232	49,194	49,163	49,114	49,059	49,009
PRODUCTION AMOUNT (MT)	-	-	-	-	-	-	-	-	-	-
BEGINNING INVENTORY	-	-	-	-	-	-	-	-	-	-
ENDING INVENTORY	-	-	-	-	-	-	-	-	-	-
COST OF GOODS SOLD	34,298	41,810	49,316	49,273	49,232	49,194	49,163	49,114	49,059	49,009

Table 5-13 Cost of Goods Manufactured and Sold
Electrolysis Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
VARIABLE COST										
Raw Materials	1,432	1,790	2,148	2,148	2,148	2,148	2,148	2,148	2,148	2,148
By-products (credit)	(143)	(178)	(214)	(214)	(214)	(214)	(214)	(214)	(214)	(214)
Utilities	5,240	6,550	7,860	7,860	7,860	7,860	7,860	7,860	7,860	7,860
Sub-total	6,529	8,162	9,794	9,794	9,794	9,794	9,794	9,794	9,794	9,794
FIXED COST										
Direct Labor	361	361	361	361	361	361	361	361	361	361
Depreciation	2,527	2,527	2,527	2,527	2,527	2,527	2,527	2,527	2,527	2,527
Repair & Maintenance	225	451	676	676	676	676	676	676	676	676
Real Property Tax & Insurance	367	331	296	260	225	190	154	119	84	48
Plant Overhead	72	72	72	72	72	72	72	72	72	72
Sub-total	3,552	3,742	3,932	3,896	3,861	3,826	3,790	3,755	3,720	3,684
CONTINGENCY	228	268	307	309	310	313	316	316	315	316
TOTAL PRODUCTION COST	10,309	12,172	14,033	13,999	13,965	13,933	13,900	13,865	13,829	13,794
PRODUCTION AMOUNT (MT)	-	-	-	-	-	-	-	-	-	-
BEGINNING INVENTORY	-	-	-	-	-	-	-	-	-	-
ENDING INVENTORY	-	-	-	-	-	-	-	-	-	-
COST OF GOODS SOLD	10,309	12,172	14,033	13,999	13,965	13,933	13,900	13,865	13,829	13,794

Table 5-14 Cost of Goods Manufactured and Sold

VCM Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1909
VARIABLE COST										
Raw Materials	13,673	17,091	20,509	20,509	20,509	20,509	20,509	20,509	20,509	20,509
By-products (credit)	-	-	-	-	-	-	-	-	-	-
Utilities	1,366	1,708	2,049	2,049	2,049	2,049	2,049	2,049	2,049	2,049
Sub-total	15,039	18,799	22,558	22,558	22,558	22,558	22,558	22,558	22,558	22,558
FIXED COST										
Direct Labor	364	364	364	364	364	364	364	364	364	364
Depreciation	2,401	2,401	2,401	2,401	2,401	2,401	2,401	2,401	2,401	2,401
Repair & Maintenance	276	551	827	827	827	827	827	827	827	827
Real Property Tax & Insurance	359	325	292	258	225	191	157	124	90	57
Plant Overhead	73	73	73	73	73	73	73	73	73	73
Sub-total	3,473	3,714	3,957	3,923	3,890	3,856	3,822	3,789	3,755	3,722
CONTINGENCY	402	491	576	581	586	593	602	604	603	605
TOTAL PRODUCTION COST	18,914	23,004	27,091	27,062	27,034	27,007	26,982	26,951	26,916	26,885
PRODUCTION AMOUNT (MT)	-	-	-	-	-	-	-	-	-	-
BEGINNING INVENTORY	-	-	-	-	-	-	-	-	-	-
ENDING INVENTORY	-	-	-	-	-	-	-	-	-	-
COST OF GOODS SOLD	18,914	23,004	27,091	27,062	27,034	27,007	26,982	26,951	26,916	26,885

Table 5-15 Cost of Goods Manufactured and Sold

PP Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
VARIABLE COST										
Raw Materials	29,611	37,014	44,416	44,416	44,416	44,416	44,416	44,416	44,416	44,416
By-products (credit)	-	-	-	-	-	-	-	-	-	-
Utilities	8,029	10,036	12,043	12,043	12,043	12,043	12,043	12,043	12,043	12,043
Sub-total	37,640	47,050	56,459	56,459	56,459	56,459	56,459	56,459	56,459	56,459
FIXED COST										
Direct Labor	492	492	492	492	492	492	492	492	492	492
Depreciation	6,833	6,833	6,833	6,833	6,833	6,833	6,833	6,833	6,833	6,833
Repair & Maintenance	604	1,207	1,811	1,811	1,811	1,811	1,811	1,811	1,811	1,811
Real Property Tax & Insurance	998	902	807	711	615	520	424	328	233	137
Plant Overhead	98	98	98	98	98	98	98	98	98	98
Sub-total	9,025	9,532	10,041	9,945	9,849	9,754	9,658	9,562	9,467	9,371
CONTINGENCY	1,019	1,238	1,446	1,464	1,484	1,507	1,536	1,542	1,540	1,546
TOTAL PRODUCTION COST	47,684	57,820	67,946	67,868	67,792	67,720	67,653	67,653	67,466	67,376
PRODUCTION AMOUNT (MT)	-	-	-	-	-	-	-	-	-	-
BEGINNING INVENTORY	-	-	-	-	-	-	-	-	-	-
ENDING INVENTORY	-	-	-	-	-	-	-	-	-	-
COST OF GOODS SOLD	47,684	57,820	67,946	67,868	67,792	67,720	67,653	67,653	67,466	67,376

5-2-4 Projected Operating Expenses

The projected operating expenses are as shown in Tables 5-16 through 5-22:

Table 5-16 Ethylene plant
Table 5-17 Utility center
Table 5-18 LDPE plant
Table 5-19 HDPE plant
Table 5-20 Electrolysis plant
Table 5-21 VCM plant
Table 5-22 PP plant

Table 5-16 Projected Operating Expense
Ethylene Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
GENERAL ADMINISTRATIVE EXPENSE	2,272	2,681	3,090	3,083	3,076	3,069	3,062	3,054	3,047	3,040
SELLING EXPENSE (PACKAGING, etc.)	-	-	-	-	-	-	-	-	-	-
AMORTIZATION										
License & Knowhow Fee	192	192	192	192	192	192	192	192	192	192
Pre-operating Expense										
Organizational Expense	447	447	447	447	447	447	447	447	447	447
Start-up Expense	603	603	603	603	603	603	603	603	603	603
Financial Charge										
L/G Fee	874	874	874	874	874	874	874	874	874	874
Interest during Construction	2,574	2,574	2,574	2,574	2,574	2,574	2,574	2,574	2,574	2,574
SALES TAX	-	-	-	-	-	-	-	-	-	-
TOTAL	6,962	7,371	7,780	7,773	7,766	7,759	7,752	7,744	7,737	7,730

Table 5-17 Projected Operating Expense
Utility Center

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
GENERAL ADMINISTRATIVE EXPENSE	964	1,140	1,325	1,323	1,320	1,318	1,315	1,313	1,310	1,308
SELLING EXPENSE (PACKAGING, etc.)	-	-	-	-	-	-	-	-	-	-
AMORTIZATION										
License & Knowhow Fee	-	-	-	-	-	-	-	-	-	-
Pre-operating Expense										
Organizational Expense	212	212	212	212	212	212	212	212	212	212
Start-up Expense	153	153	153	153	153	153	153	153	153	153
Financial Charge										
L/G Fee	326	326	326	326	326	326	326	326	326	326
Interest during Construction	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000
SALES TAX	-	-	-	-	-	-	-	-	-	-
TOTAL	2,655	2,831	3,016	3,014	3,011	3,009	3,006	3,004	3,001	2,999

Table 5-18 Projected Operating Expense
LDPE Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
GENERAL ADMINISTRATIVE EXPENSE	1,918	2,267	2,599	2,629	2,663	2,708	2,765	2,776	2,772	2,782
SELLING EXPENSE (PACKAGING, etc.)	990	1,237	1,485	1,485	1,485	1,485	1,485	1,485	1,485	1,485
AMORTIZATION										
License & Knowhow Fee	596	596	596	596	596	596	596	596	596	596
Pre-operating Expense										
Organizational Expense	250	250	250	250	250	250	250	250	250	250
Start-up Expense	363	363	363	363	363	363	363	363	363	363
Financial Charge										
L/G Fee	489	489	489	489	489	489	489	489	489	489
Interest during Construction	1,462	1,462	1,462	1,462	1,462	1,462	1,462	1,462	1,462	1,462
SALES TAX	670	1,616	1,904	3,014	4,259	5,894	7,939	8,406	8,406	8,872
TOTAL	6,738	8,280	9,148	10,288	11,567	13,247	15,349	15,827	15,823	16,299

Table 5-19 Projected Operating Expense
HDPE Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
GENERAL ADMINISTRATIVE EXPENSE	1,096	1,332	1,557	1,573	1,591	1,614	1,648	1,656	1,654	1,660
SELLING EXPENSE (PACKAGING, etc.)	810	810	810	810	810	810	810	810	810	810
AMORTIZATION										
License & Knowhow Fee	506	506	506	506	506	506	506	506	506	506
Pre-operating Expense										
Organizational Expense	120	120	120	120	120	120	120	120	120	120
Start-up Expense	223	223	223	223	223	223	223	223	223	223
Financial Charge										
L/G Fee	238	238	238	238	238	238	238	238	238	238
Interest during Construction	727	727	727	727	727	727	727	727	727	727
SALES TAX	362	879	1,040	1,627	2,272	3,106	4,290	4,585	4,585	4,840
TOTAL	4,082	4,835	5,221	5,824	6,487	7,344	8,562	8,865	8,863	9,124

Table 5-20 Projected Operating Expense
Electrolysis Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
GENERAL ADMINISTRATIVE EXPENSE	333	391	447	449	452	455	460	460	459	460
SELLING EXPENSE (PACKAGING, etc.)	-	-	-	-	-	-	-	-	-	-
AMORTIZATION										
License & Knowhow Fee	181	181	181	181	181	181	181	181	181	181
Pre-operating Expense										
Organizational Expense	74	74	74	74	74	74	74	74	74	74
Start-up Expense	80	80	80	80	80	80	80	80	80	80
Financial Charge										
L/G Fee	145	145	145	145	145	145	145	145	145	145
Interest during Construction	443	443	443	443	443	443	443	443	443	443
SALES TAX	81	202	243	364	486	631	825	874	874	922
TOTAL	1,337	1,516	1,613	1,736	1,861	2,009	2,208	2,257	2,256	2,305

Table 5-21 Projected Operating Expense
VCM Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
GENERAL ADMINISTRATIVE EXPENSE	586	715	838	846	854	864	877	880	879	881
SELLING EXPENSE (PACKAGING, etc.)	-	-	-	-	-	-	-	-	-	-
AMORTIZATION										
*License & Knowhow Fee	100	100	100	100	100	100	100	100	100	100
Pre-operating Expense										
Organizational Expense	68	68	68	68	68	68	68	68	68	68
Start-up Expense	134	134	134	134	134	134	134	134	134	134
Financial Charge										
L/G Fee	133	133	133	133	133	133	133	133	133	133
Interest during Construction	389	389	389	389	389	389	389	389	389	389
SALES TAX	200	499	599	899	1,198	1,558	2,037	2,157	2,157	2,277
TOTAL :	1,610	2,038	2,261	2,569	2,876	3,246	3,738	3,861	3,860	3,982

* Including Catalyst and Chemicals

Table 5-22 Projected Operating Expense
PP Plant

(1,000 US$)

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
GENERAL ADMINISTRATIVE EXPENSE	1,528	1,858	2,169	2,196	2,225	2,260	2,304	2,313	2,310	2,319
SELLING EXPENSE (PACKAGING, etc.)	810	1,012	1,215	1,215	1,215	1,215	1,215	1,215	1,215	1,215
AMORTIZATION										
License & Knowhow Fee	764	764	764	764	764	764	764	764	764	764
Pre-operating Expense										
Organizational Expense	202	202	202	202	202	202	202	202	202	202
Start-up Expense	289	289	289	289	289	289	289	289	289	289
Financial Charge										
L/G Fee	398	398	398	398	398	398	398	398	398	398
Interest during Construction	1,209	1,209	1,209	1,209	1,209	1,209	1,209	1,209	1,209	1,209
SALES TAX	608	1,468	1,729	2,710	3,791	5,055	6,610	6,999	6,999	7,388
TOTAL	5,808	7,200	7,975	8,983	10,093	11,392	12,991	13,389	13,386	13,784

5-2-5 Results of Financial Analysis

Results of financial analysis based on the BOI's standard is as shown in Tables 5-23 through 5-39 as follows:

Table 5-23 Results of Financial Analysis

Discounted Cash Flow on Total Capital Requirement

Table 5-24	Whole complex
Table 5-25	Ethylene plant
Table 5-26	Utility center
Table 5-27	LDPE plant
Table 5-28	HDPE plant
Table 5-29	Electrolysis plant
Table 5-30	VCM plant
Table 5-31	PP plant

Discounted Cash Flow on Paid-in Capital

Table 5-32	Whole complex
Table 5-33	Ethylene plant
Table 5-34	Utility center
Table 5-35	LDPE plant
Table 5-36	HDPE plant
Table 5-37	Electrolysis plant
Table 5-38	VCM plant
Table 5-39	PP plant

Table 5-23 Results of Financial Analysis

	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
A. PROFITABILITY										
a. ROI on Total Assets (%)	4.91	7.05	9.47	10.44	11.48	12.04	11.73	11.21	10.76	10.21
b. ROI on Stockholders Equity (%)	0.30	9.64	15.31	16.19	16.69	16.35	14.79	13.28	12.13	11.09
c. Net Profit Margin on Sales x Turnover of Investment (%)	7.55	10.85	14.56	16.06	17.66	18.52	18.05	17.25	16.55	15.71
B. LIQUIDITY RATIOS										
a. Current Ratio (%)	75.46	103.31	139.38	186.94	246.66	301.97	361.89	423.82	486.00	796.08
b. Quick Ratio (%)	75.37	103.21	139.28	186.84	246.55	301.87	361.79	423.72	485.90	795.94
c. Day's Sales in Receivables (Days)	67	67	67	67	67	67	67	67	67	67
C. SOLVENCY RATIOS										
a. Debt-Equity Ratio	71:29	68:32	62:38	55:45	46:54	37:63	30:70	23:77	16:84	10:90
b. Debt-Service Coverage on Contractual Obligations (%)	69.45	85.27	100.33	123.39	158.72	212.67	226.90	239.37	267.01	284.43

Table 5-24 Discounted Cash Flow on Total Investment
Whole Complex

(1,000 US$)

	Total Capital Require-ment	Cash In-flow			Present Value at 16.0 %		
		Net Cash Flow	*Repayment	Total	Present Factor	Invest-ment	Cash In-flow
Pre-operation	727,013	40,037	26,030	66,067	1.00000	660,946	
1980		(7,556)	92,382	84,826	0.86207		73,126
1981		38,114	87,952	126,066	0.74316		93,687
1982		48,931	83,523	132,454	0.64066		84,858
1983		57,196	79,092	136,288	0.55229		75,270
1984		71,197	74,661	145,858	0.47611		69,444
1985		94,341	70,231	164,572	0.41044		67,547
1986		96,632	65,724	162,356	0.35383		57,446
1987		99,658	61,371	161,029	0.30503		49,119
1988		103,376	56,941	160,317	0.26295		42,155
1989		159,857	52,512	212,369	0.22668		48,140
Total	727,013					660,946	660,792

* Repayment of Long Term Loan and Financial Charges Payment

Table 5-25 Discounted Cash Flow on Total Investment
Ethylene Plant

(1,000 US$)

	Total Capital Require-ment	Cash In-flow			Present Value at 15.1 %		
		Net Cash Flow	*Repayment	Total	Present Factor	Invest-ment	Cash In-flow
Pre-operation	236,145	12,551	8,740	21,291	1.00000	214,854	
1980		(2,491)	30,074	27,583	0.86881		23,964
1981		14,254	28,632	42,886	0.75483		32,372
1982		18,131	27,189	45,320	0.65580		29,721
1983		18,372	25,747	44,119	0.56977		25,138
1984		19,960	24,305	44,265	0.49502		21,912
1985		25,731	22,863	48,594	0.43008		20,899
1986		26,568	21,420	47,988	0.37366		17,931
1987		27,667	19,979	47,646	0.32464		15,468
1988		28,766	18,536	47,302	0.28205		13,342
1989		37,287	17,095	54,382	0.24505		13,326
Total	236,145					214,854	214,073

* Repayment of Long Term Loan and Financial Charges Payment

Table 5-26 Discounted Cash Flow on Total Investment
Utility Center

(1,000 US$)

	Total Capital Requirement	Cash In-flow			Present Value at 11.0 %		
		Net Cash Flow	*Repayment	Total	Present Factor	Investment	Cash In-flow
Pre-operation	110,933	4,873	3,260	8,133	1.00000	102,800	
1980		(4,745)	14,264	9,519	0.90090		8,576
1981		1,851	13,580	15,431	0.81162		12,524
1982		4,296	12,897	17,193	0.73119		12,571
1983		3,000	12,213	15,213	0.65873		10,021
1984		5,013	11,528	16,541	0.59345		9,816
1985		6,983	10,844	17,827	0.53464		9,531
1986		7,384	10,160	17,544	0.48166		8,450
1987		7,884	9,476	17,360	0.43393		7,533
1988		8,385	8,792	17,177	0.39092		6,715
1989		38,936	8,108	47,044	0.35218		16,568
Total	110,933					102,800	102,305

* Repayment of Long Term Loan and
Financial Charges Payment

Table 5-27 Discounted Cash Flow on Total Investment
LDPE Plant

(1,000 US$)

	Total Capital Requirement	Cash In-flow			Present Value at 18.6 %		
		Net Cash Flow	*Repayment	Total	Present Factor	Investment	Cash In-flow
Pre-operation	131,637	6,832	4,890	11,722	1.00000	119,915	
1980		(1,951)	16,785	14,834	0.84317		12,508
1981		6,995	15,980	22,975	0.71094		16,334
1982		8,968	15,175	24,143	0.59944		14,472
1983		12,918	14,371	27,289	0.50543		13,793
1984		17,485	13,566	31,051	0.42616		13,233
1985		24,629	12,761	37,390	0.35933		13,435
1986		25,381	11,956	37,337	0.30298		11,312
1987		25,578	11,151	36,729	0.25546		9,383
1988		26,359	10,346	36,705	0.21540		7,906
1989		31,503	9,541	41,044	0.18162		7,454
Total	131,637					119,915	119,830

* Repayment of Long Term Loan and
Financial Charges Payment

Table 5-28 Discounted Cash Flow on Total Investment HDPE Plant

(1,000 US$)

	Total Capital Requirement	Cash In-flow			Present Value at 14.1 %		
		Net Cash Flow	*Repayment	Total	Present Factor	Investment	Cash In-flow
Pre-operation	64,245	4,000	2,380	6,380	1.00000	57,865	
1980		(2,399)	8,103	5,704	0.87642		4,999
1981		499	7,714	8,213	0.76812		6,309
1982		808	7,326	8,134	0.67320		5,476
1983		3,363	6,936	10,299	0.59001		6,077
1984		5,563	6,548	12,111	0.51710		6,263
1985		8,499	6,159	14,658	0.45320		6,643
1986		10,341	5,771	16,112	0.39719		6,400
1987		10,609	5,382	15,991	0.34811		5,567
1988		10,880	4,994	15,874	0.30509		4,843
1989		14,427	4,605	19,032	0.26739		5,089
Total	64,245					57,865	57,666

* Repayment of Long Term Loan and Financial Charges Payment

Table 5-29 Discounted Cash Flow on Total Investment Electrolysis Plant

(1,000 US$)

	Total Capital Requirement	Cash In-flow			Present Value at 14.1 %		
		Net Cash Flow	*Repayment	Total	Present Factor	Investment	Cash In-flow
Pre-operation	38,930	1,963	1,450	3,413	1.00000	35,517	
1980		(570)	4,957	4,387	0.87642		3,845
1981		2,131	4,720	6,851	0.76812		5,262
1982		2,883	4,482	7,365	0.67320		4,958
1983		2,772	4,244	7,016	0.59001		4,140
1984		2,949	4,006	6,955	0.51710		3,596
1985		3,788	3,768	7,556	0.45320		3,424
1986		3,773	3,531	7,304	0.39719		2,901
1987		3,972	3,293	7,265	0.34811		2,529
1988		4,169	3,055	7,224	0.30509		2,204
1989		6,660	2,818	9,478	0.26739		2,534
Total	38,930					35,517	35,393

* Repayment of Long Term Loan and Financial Charges Payment

Table 5-30 Discounted Cash Flow on Total Investment
VCM Plant

(1,000 US$)

	Total Capital Require-ment	Cash In-flow			Present Value at 11.9 %		
		Net Cash Flow	*Repayment	Total	Present Factor	Invest-ment	Cash In-flow
Pre-operation	36,239	2,124	1,330	3,454	1.00000	32,785	
1980		(740)	4,588	3,848	0.89366		3,439
1981		1,640	4,368	6,008	0.79862		4,798
1982		2,324	4,148	6,472	0.71369		4,619
1983		2,052	3,928	5,980	0.63779		3,814
1984		2,097	3,708	5,805	0.56997		3,309
1985		2,785	3,488	6,273	0.50935		3,195
1986		2,614	3,191	5,805	0.45519		2,642
1987		2,724	3,048	5,772	0.40678		2,348
1988		2,962	2,828	5,790	0.36352		2,105
1989		4,866	2,609	7,475	0.32486		2,428
Total	36,239					32,785	32,697

* Repayment of Long Term Loan and Financial Charges Payment

Table 5-31 Discounted Cash Flow on Total Investment
PP Plant

(1,000 US$)

	Total Capital Require-ment	Cash In-flow			Present Value at 23.0 %		
		Net Cash Flow	*Repayment	Total	Present Factor	Invest-ment	Cash In-flow
Pre-operation	108,884	7,694	3,980	11,674	1.00000	97,210	
1980		5,340	13,611	18,951	0.81301		15,407
1981		10,744	12,958	23,702	0.66098		15,667
1982		11,521	12,306	23,827	0.53738		12,804
1983		14,719	11,653	26,372	0.43690		11,522
1984		18,130	11,000	29,130	0.35520		10,347
1985		21,926	10,348	32,274	0.28878		9,320
1986		20,571	9,695	30,266	0.23478		7,106
1987		21,224	9,042	30,266	0.19088		5,777
1988		21,855	8,390	30,245	0.15519		4,694
1989		26,178	7,736	33,914	0.12617		4,279
Total	108,884					97,210	96,923

* Repayment of Long Term Loan and Financial Charges Payment

Table 5-32 Discounted Cash Flow on Paid-in Capital
Whole Complex

(1,000 US$)

	Paid-in Capital	Net Cash In-flow	Present Value at 22.5%		
			Present Factor	Paid-in Capital	Cash In-flow
Pre-operation	206,064		1.00000	206,064	
1980		(7,556)	0.81633		(6,168)
1981		38,114	0.66639		25,399
1982		48,931	0.54399		26,618
1983		57,196	0.44407		25,399
1984		71,197	0.36251		25,810
1985		94,341	0.29593		27,918
1986		96,632	0.24157		23,343
1987		99,658	0.19720		19,653
1988		103,376	0.16098		16,641
1989		159,857	0.13141		21,007
Total	206,064			206,064	205,620

Table 5-33 Discounted Cash Flow on Paid-in Capital
Ethylene Plant

(1,000 US$)

	Paid-in Capital	Net Cash In-flow	Present Value at 20.7%		
			Present Factor	Paid-in Capital	Cash In-flow
Pre-operation	67,074		1.00000	67,074	
1980		(2,491)	0.82850		(2,064)
1981		14,254	0.68641		9,784
1982		18,131	0.56869		10,311
1983		18,372	0.47116		8,656
1984		19,960	0.39036		7,792
1985		25,731	0.32341		8,322
1986		26,568	0.26795		7,119
1987		27,667	0.22199		6,142
1988		28,766	0.18392		5,291
1989		37,287	0.15238		5,682
Total	67,074			67,074	67,035

Table 5-34 Discounted Cash Flow on Paid-in Capital
Utility Center

(1,000 US$)

	Paid-in Capital	Net Cash In-flow	Present Value at 11.6%		
			Present Factor	Paid-in Capital	Cash In-flow
Pre-operation	31,820		1.00000	31,820	
1980		(4,745)	0.89606		(4,252)
1981		1,851	0.80292		1,486
1982		4,296	0.71946		3,091
1983		3,000	0.64468		1,934
1984		5,013	0.57767		2,896
1985		6,983	0.51762		3,615
1986		7,384	0.46382		3,425
1987		7,884	0.41561		3,277
1988		8,385	0.37241		3,123
1989		38,936	0.33370		12,993
Total	31,820			31,820	31,588

Table 5-35 Discounted Cash Flow on Paid-in Capital
LDPE Plant

(1,000 US$)

	Paid-in Capital	Net Cash In-flow	Present Value at 27.3%		
			Present Factor	Paid-in Capital	Cash In-flow
Pre-operation	37,445		1.00000	37,445	
1980		(1,951)	0.78555		(1,533)
1981		6,995	0.61708		4,316
1982		8,968	0.48475		4,347
1983		12,918	0.38079		4,919
1984		17,485	0.29913		5,230
1985		24,629	0.23498		5,787
1986		25,381	0.18459		4,685
1987		25,578	0.14500		3,709
1988		26,359	0.11391		3,003
1989		31,503	0.08948		2,819
Total	37,445			37,445	37,282

Table 5-36 Discounted Cash Flow on Paid-in Capital
HDPE Plant

(1,000 US$)

	Paid-in Capital	Net Cash In-flow	Present Value at 17.5%		
			Present Factor	Paid-in Capital	Cash In-flow
Pre-operation	18,075		1.00000	18,075	
1980		(2,399)	0.85106		(2,042)
1981		499	0.72431		361
1982		808	0.61643		498
1983		3,363	0.52462		1,764
1984		5,563	0.44649		2,484
1985		8,499	0.37999		3,230
1986		10,341	0.32340		3,344
1987		10,609	0.27523		2,920
1988		10,880	0.23424		2,549
1989		14,427	0.19935		2,876
Total	18,075			18,075	17,984

Table 5-37 Discounted Cash Flow on Paid-in Capital
Electrolysis Plant

(1,000 US$)

	Paid-in Capital	Net Cash In-flow	Present Value at 18.6%		
			Present Factor	Paid-in Capital	Cash In-flow
Pre-operation	11,065		1.00000	11,065	
1980		(570)	0.84317		(481)
1981		2,131	0.71094		1,515
1982		2,883	0.59944		1,728
1983		2,772	0.50543		1,401
1984		2,949	0.42616		1,257
1985		3,788	0.35933		1,361
1986		3,773	0.30298		1,143
1987		3,972	0.25546		1,015
1988		4,169	0.21540		898
1989		6,660	0.18162		1,210
Total	11,065			11,065	11,047

Table 5-38 Discounted Cash Flow on Paid-in Capital
VCM Plant

(1,000 US$)

	Paid-in Capital	Net Cash In-flow	Present Value at 13.6%		
			Present Factor	Paid-in Capital	Cash In-flow
Pre-operation	10,235		1.00000	10,235	
1980		(740)	0.88028		(651)
1981		1,640	0.77490		1,271
1982		2,324	0.68213		1,585
1983		2,052	0.60046		1,232
1984		2,097	0.52858		1,108
1985		2,785	0.46530		1,296
1986		2,614	0.40959		1,071
1987		2,724	0.36056		982
1988		2,962	0.31739		940
1989		4,866	0.27939		1,360
Total	10,235			10,235	10,194

Table 5-39 Discounted Cash Flow on Paid-in Capital
PP Plant

(1,000 US$)

	Paid-in Capital	Net Cash In-flow	Present Value at 38.5%		
			Present Factor	Paid-in Capital	Cash In-flow
Pre-operation	30,350		1.00000	30,350	
1980		5,340	0.72202		3,856
1981		10,744	0.52132		5,601
1982		11,521	0.37640		4,337
1983		14,719	0.27177		4,000
1984		18,130	0.19622		3,557
1985		21,926	0.14168		3,106
1986		20,571	0.10229		2,104
1987		21,224	0.07386		1,568
1988		21,855	0.05333		1,166
1989		26,178	0.03850		1,008
Total	30,350			30,350	30,303

5-2-6 Measurement of Economic Factors

The "Measurement of Economic Factors" based on the BOI's criteria are as shown in Table 5-40.

Table 5-40 Measurement of Economic Factors

	1	2	3	4	5	6	7	8	9	10	Average
(a) Value Added Coefficient (%)	42.17	42.17	41.93	43.34	44.86	46.23	46.95	47.01	47.01	47.01	45.12
(b) Usage of Indigenous Raw Materials (%)	47.52	47.51	47.51	47.51	47.51	47.51	47.51	47.51	47.51	47.51	47.51
(c) Fixed Assets per Worker (1,000US$)	457.71	412.17	366.63	321.09	275.55	229.89	184.48	138.94	93.41	47.87	252.78
(d) Fixed Assets per Peso Wage/ Salary (US$)	114.10	102.75	91.40	80.05	68.69	57.30	45.99	34.64	23.29	11.93	63.01
(e) Fixed Assets/Value Added (%)	3.15	2.33	1.76	1.45	1.17	0.92	0.72	0.54	0.36	0.19	1.12
(f) Foreign Exchange Benefit-Cost Ratio (%)	228.33	256.89	281.74	294.51	308.84	323.26	334.82	342.32	349.93	357.89	307.67
(g) $\frac{\text{Foreign Exchange Benefit}}{\text{Imported Raw Materials}}$ (%)	494.35	485.98	481.67	493.62	507.27	520.12	527.26	527.83	527.83	527.83	510.32
* Sales (1,000 US$)	359,067	441,242	524,796	537,819	552,689	566,684	574,470	575,090	575,090	575,090	
* Value Added (1,000 US$)	153,109	186,054	220,062	233,085	247,955	261,950	269,736	270,356	270,356	270,356	

Note: Formula

(a) $\frac{\text{Value of Output} - (\text{Raw Materials} + \text{Utilities} + \text{Supplies})}{\text{Value of Output}}$

(b) $\frac{\text{Cost of Local Raw Materials}}{\text{Total Raw Material Cost}}$

(c) $\frac{\text{Fixed Assets Excluding Land}}{\text{Number of Workers}}$

(d) $\frac{\text{Fixed Assets Excluding Land}}{\text{Total Payroll}}$

(e) $\frac{\text{Fixed Assets Excluding Land}}{\text{Value of Output} - (\text{Raw Materials} + \text{Utilities} + \text{Supplies})}$

(f) $\frac{\text{Gross Foreign Exchange Savings/Earnings}}{\text{Foreign Exchange Outflows}}$

5-2-7 Foreign Currency Balance

The expected balance of foreign exchange flow is as shown in Table 5-41.

Table 5-41 Summary of Foreign Currency Balance

(1,000 US$)

	Pre-Operation	1980	1981	1982	1983	1984	1985	1986	1987	1988	1989
OUT-FLOW											
Raw Materials	5,434	72,635	90,794	108,953	108,953	108,953	108,953	108,953	108,953	108,953	108,953
Spare Parts	17,640	2,948	5,897	8,845	8,845	8,845	8,845	8,845	8,845	8,845	8,845
*Others	-	1,630	1,993	2,356	2,356	2,356	2,356	2,356	2,356	2,356	2,356
**Plant Equipment	435,070	-	-	-	-	-	-	-	-	-	-
Interest during Construction	78,040	-	-	-	-	-	-	-	-	-	-
Interest on Long Term Loan	-	36,542	32,886	29,233	25,580	21,924	18,270	14,541	10,962	7,308	3,654
Total Out-flow	536,184	113,755	131,570	149,387	145,734	142,078	138,424	134,695	131,116	127,462	123,808
IN-FLOW											
Export of Products	-	29,132	45,961	61,047	44,977	26,638	10,051	919	-	-	-
Saving by Import Substitution	-	134,561	155,424	178,620	200,136	224,692	246,926	259,171	260,400	260,400	260,400
Total In-flow	-	163,693	201,385	239,667	245,113	251,330	256,977	260,090	260,400	260,400	260,400
BALANCE											
Balance	-536,184	49,938	69,815	90,280	99,379	109,252	118,553	125,395	129,284	132,938	136,592
Accumurated Balance	-536,184	-486,246	-416,431	-326,151	-226,772	-117,520	1,033	126,428	255,712	388,650	525,242

* Contingency on Raw Materials and Spare Parts

** Including Equipment, Construction Materials, Licence and Knowhow Fee, Catalyst and Chemicals

5-2-8 Others

Repayment schedule for the long-term loan for the total complex is as shown in Table 5-42.

Table 5-42 Repayment Schedule for Long-term Loan

Total Loan Amount: US$480,810

	Outstanding Balance of Long Term Loan	Repayment			Guarantee Fee	Total Payment
		A. Principal	B. Interest	A + B		
	480,810					
1980	432,729	48,081	36,542	84,623	7,759	92,382
1981	384,648	48,081	32,886	80,967	6,985	87,952
1982	336,567	48,081	29,233	77,314	6,209	83,523
1983	288,486	48,081	25,580	73,661	5,431	79,092
1984	240,405	48,081	21,924	70,005	4,656	74,661
1985	192,324	48,081	18,270	66,351	3,880	70,231
1986	144,243	48,081	14,541	62,622	3,102	65,724
1987	96,162	48,081	10,962	59,043	2,328	61,371
1988	48,081	48,081	7,308	55,389	1,552	56,941
1989	0	48,081	3,654	51,735	777	52,512
Total		480,810	200,900	681,710	42,679	724,389

CHAPTER 6. STUDIES ON PROCESS ROUTES, POSSIBLE IMPROVEMENT IN PRODUCTION COST

CHAPTER 6. STUDIES ON PROCESS ROUTES, POSSIBLE IMPROVEMENT IN PRODUCTION COST

CONTENTS

TABLES AND ILLUSTRATIONS

6. Studies on Process Routes, Possible Improvement in Production Cost

6-1 Comparison of Process Routes and Sites for VCM Production

6-1-1 Introduction

The four possible sources of VCM for the production of PVC in the Philippines are as follows:

(a) Import of VCM(Existing)

(b) Acetylene route: VCM production through carbide basis acetylene(Existing)

(c) Ethylene route: VCM production through oxychlorination of EDC domestically produced from ethylene, which is produced from domestic naphtha cracker

(d) EDC route: VCM production through oxychlorination of imported EDC with domestically produced ethylene.

The selection of VCM sources will form the frame work of the petrochemical complex. The production costs of the above-mentioned process routes (a), (b), (c), and (d) are investigated for each of the following candidate sites to determine relative advantages and disadvantages.

Based on these investigation, the candidate sites for VCM plants were nominated on the potential for actualization and are as follows:

(1) Iligan City, Mindanao

This plant site is applicable to (b) of the above process routes where carbide can be produced utilizing low price hydroelectricity available from NPC Mariachristina, and limestone and charcoal are avilable domestically and by importation from foreign countries respectively, as presently practiced by the Mariachristina Carbide Corporation.

In the Iligan area, the Mariachristina Carbide Corp. and the Mabuhay Vinyl Corp. jointly produce VCM. The Mariachristina Carbide Corp. supplies acetylene to the Mabuhay Vinyl Corp. The latter produces VCM from acetylene utilizing an electrolysis plant for the supply of chlorine.

At the present time, the production capacity of VCM by the above-mentioned companies are 9,000 t/y, and the Mabuhay Vinyl Corp. has an extra production capacity of PVC up to 22,000t/y in total using imported VCM. The Mabuhay Vinyl Corp. intends to expand their existing production line based on the acetylene route to twice the present capacity.

(2) Cavite City, Luzon

This plant site applies to process routes (a), (c) and (d). Cavite has been nominated as tentative project site for a petrochemical complex due to its proximity to the major markets in the Manila area. This is highly advantageous in terms of project costs. In terms of process routes (c) and (d), if a petrochemical complex is established in Cavite, VCM and PVC production can be integrated with ethylene supplied by naphtha cracker. One disadvantage is the higher electric power costs for the electrolysis plant.

Philippine Vinyl Consotium, Inc. is now constructing a PVC plant with a capacity of 25,000 t/y on the basis of imported VCM.

6-1-2 Comparison of Ethylene Route - Acetylene Route and Project Sites for VCM Production

A preliminary comparison of the ethylene route and the acetylene route VCM production costs has to be reevaluated due to the oil crisis which has drastically raised petroleum costs. Another important factor to be considered then, is the low electricity costs in the case of Iligan. On the basis of the above comparison, tentative supply sources are studied by taking into account the location factors of Iligan and Manila area. The flow of each process route is shown in Figs. 6-1 and 6-2.

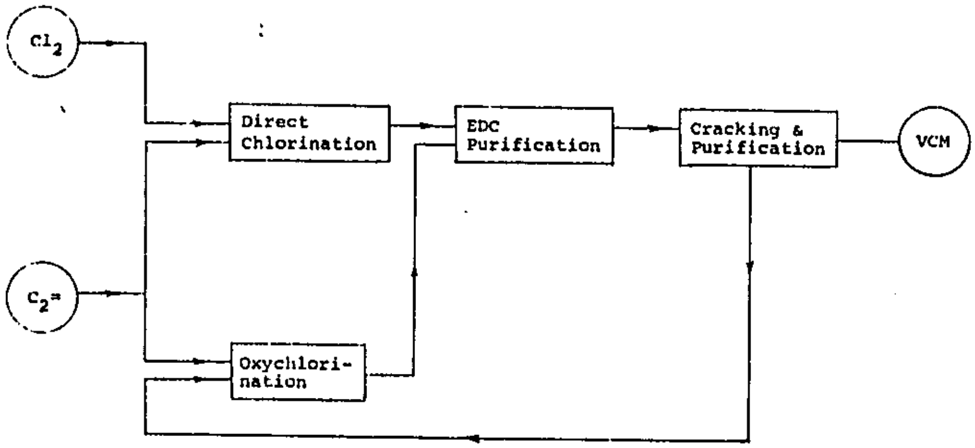


Fig. 6-1 Ethylene Route Process Flow Block Diagram of VCM Production

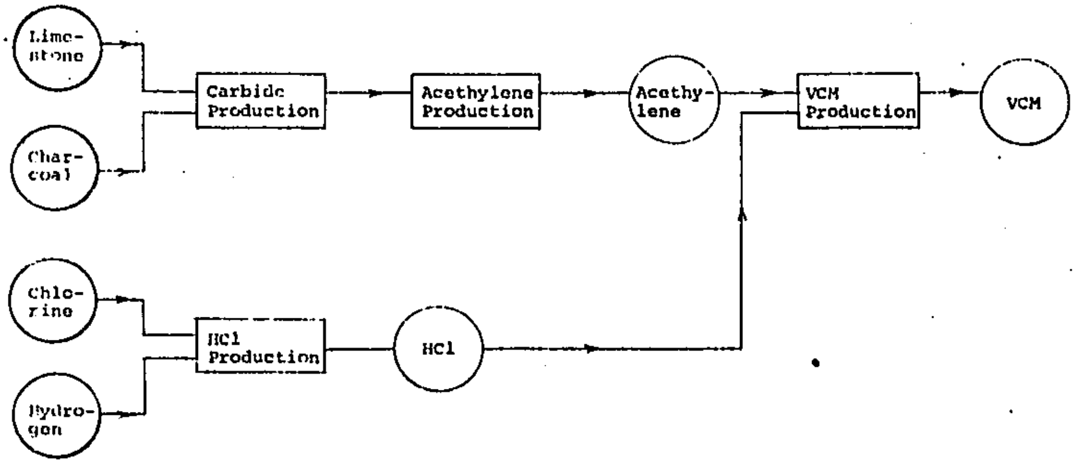


Fig. 6-2 Acetylene Route Block Diagram of VCM Production Process Route

(1) Comparison of process routes under standard conditions

Under standard conditions as shown in Table 6-1, production costs and ex-factory prices with 15% of the internal rate of return are calculated on a trial basis for each process route. As shown in Tables 6-2 and 6-3, the production costs for each process route differ slightly. The required investment for the acetylene route is greater than that of ethylene route, however, the projected amount of profit is greater in the former case when the same level of the profit rate is applied to both cases. Consequently the ex-factory price of the acetylene route is actually greater than that of the ethylene route case.

Table 6-1 Prerequisite Conditions for Comparison of Process Route

Item	Price	Remark
Ethylene	557 US$/t	
Chlorine	110 US$/t	Ref. 6-2, Part II, (5) Sensitivity analysis, chlorine
Limestone	7 US$/t	Escalation 7%/y
Charcoal	130 US$/t	Escalation 7%/y
Electricity	3 US$/KWH	

Construction Cost of 55,000 t/y VCM Production

Ethylene Route	ISBL	17,000 x 10^3 US$ excl. pollution control fac.
	Off-site	3,000 x 10^3 US$
Acethylene Route	ISBL	41,000 x 10^3 US$ incl. carbide, HCl fac.
	Off-site	4,000 x 10^3 US$

Table 6-2 Production Cost of VCM - Ethylene Route

Capacity	VCM	55,000 t/y
Construction Cost	ISBL	$17,000 \times 10^3$ US$
	Off-Site	$3,000 \times 10^3$ US$

Production Cost

	Consumption/ Product (t/t)	Unit Price (US$/t)	Unit Cost (US$/t)
Ethylene	0.467	558	261
Chlorine	0.592	110	65
Oxygen	0.142	35	5
Misc.		8	8
Total Raw Matl.			339
Electricity	215	0.03	7
Medium Steam	0.8	11.6	9
Ind. Water	11.6	0.054	1
B.W.F.	0.036	1.01	-
N_2	39 (Nm^3/t)	0.127	5
Fuel	1.1 (MMKcal/t)	10.9	12
Total Utilities			34
Total Variable Cost			373
Labor			7
Depreciation, Amortization			98
Maintenance			
Tax & Insurance			
Interest			
Plant Overhead			2
Total Fixed Cost			107
Administrative Expense			15
Contingency			10
Total Production Cost			535
Profit & Loss Before Tax			68
Net Sales (Ex-factory price excl. sales tax)			603

Table 6-3 Production Cost of VCM - Acetylene Route

Capacity	VCM	55,000 t/y
Construction Cost	ISBL	41,000 x 10^3 US$
	Off-site	4,000 x 10^3 US$

Production Cost	Consumption/ Unit product(t/t)	Standard Unit Price (US$/t)	Unit Cost (US$/t)
Limestone	2.48	7	17
Charcoal	1.0	130	130
Chlorine	0.63	110	69
Hydrogene	0.02	261	5
Misc. Credit		-60	-60
Total Raw Matl.			161
Electricity	4,500(KWH)	0.03	135
Medium Steam	0.6	11.6	7
Cooling Water	180	0.03	5
Others		10.0	10
Total Utilities			157
Total Variable Cost			318
Labor			6
Depreciation, Amortization			
Maintenance			
Tax & Insurance			209
Interest			
Plant Overhead			6
Total Fixed Cost			221
Administrative Expense			17
Contingency			11
Total Production Cost			567
Profit & Loss before Tax			141
Net Sales (Ex-factory price excl. sales tax)			708

(2) Comparison of sites - utilization of low cost hydroelectricity

In the production of VCM through the acetylene route, low cost hydroelectricity is utilized. Thus the calibration of the price of chlorine electricity price change is considered in studying production costs. However, the price of caustic soda should be calculated at a lower rate due to the compensation of the rise of transportation costs of caustic soda from Iligan to Manila.

Calibration of the chlorine price by sites

Prerequisite basis

Electricity in Iligan

4 - 5₱/KWH@1974 x $(1.07)^5$ = 0.86 - 1.07¢/KWH = 1¢/KWH

Transportation cost of caustic soda from Iligan to Manila

US$45/t@1980

Calibration

Electricity	3,320KWH/t	3,320 x (0.0101-0.03)	=-66(US$/t)
Caustic soda	1.13t/t	1.13 x 45	=+51(US$/t)
Total calibration			-15(US$/t)

Thus, the price of chlorine in Iligan is

110 + (-15) = US$95/t

As shown in Table 6-4, the acetylene basis production costs are less than the ethylene basis production costs, while the ex-factory price of each case is nearly equal. Therefore, it becomes evident that the supply of low cost electricity is indispensable for the establishment of acetylene route VCM production.

6-1-3 Imported EDC Basis VCM Production

Large-scale petrochemical complexes are being planned in the Middle East or Australia where included the export-oriented EDC production through utilization of low cost hydrocarbon and electricity on the basis of abundant natural gas.

The EDC route process requires ethylene for VCM production as shown in Fig. 6-3. The complexity of the process differs slightly in comparison with the ethylene route process (ref. Fig. 6-1), so that the necessary investment for the process constitutes 75% of the investment for ethylene route process. Therefore, it seems unprofitable to produce EDC basis VCM, unless it is produced on a large-scale when chlorine sources are not available.

On the basis of BOI estimations, the consumption of caustic soda will reach more than 80,000 t/y in 1978 in the Philippines. To obtain the balance of consumption of caustic soda and chlorine, the consumption of chlorine should be promoted. The greatest consumer of chlorine is VCM production, so it seems unlikely to produce EDC basis VCM.

Table 6-4 Production Cost of VCM - Acetylene Route, Hydroelectricity

Capacity	VCM	55,000 t/y
Construction Cost	ISBL	41,000 x 10^3US$
	Off-Site	4,000 x 10^3US$

Production Cost

	Consumption/ Product (t/t)	Unit Price (US$/t)	Unit Cost (US$/t)
Limestone	2.48	7	17
Charcoal	1.0	130	130
Chlorine	0.63	95	60
Hydrogen	0.02	261	5
Misc. Credit		-60	-60
Total Raw Matl.			152
Electricity	4500 (KWH)	0.01	45
Midium Steam	0.6	11.6	7
Cooling Water	180	0.03	5
Others		10.0	10
Total Utilities			67
Total Variable Cost			219
Labor			6
Depreciation, Amortization			209
Maintenance			
Tax & Insurance			
Interest			
Plant Overhead			6
Total Fixed Cost			221
Administrative Expense			17
Contingency			11
Total Production Cost			468
Profit & Loss Before Tax			141
Net Sales (Ex-factory price excl. sales tax)			609

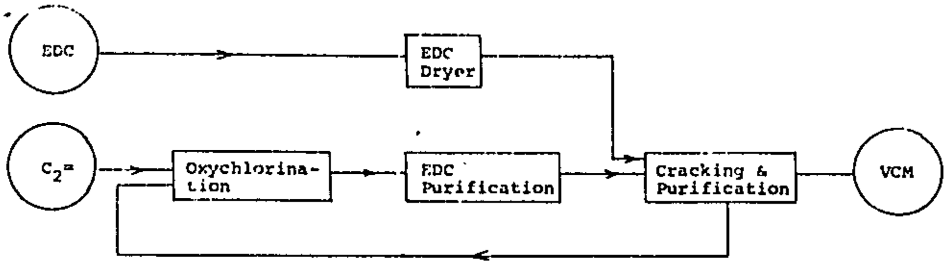


Fig. 6-3 EDC Route Process Flow Block Diagram of VCM Production

6-2 Sensitivity Analysis and Possible Improvement in Profitability and Production Cost of Each Process Plant

Generally, the sensitivity analysis is conducted throughout each process plant in the complex. For example, the rise of the price of basic raw materials, i.e., naphtha, effects the rise of the prices of intermediate products, i.e., ethylene so that the profitability of derivative plants, i.e., the LDPE plant, decrease. The effect of this factor, in the price change of intermediate products, is evaluated in the above-mentioned manner. (The level of profitability of intermediate derivative plant is specified as a 15% internal rate of return, except in the case of sensitivity analysis which is affected by the change of intermediate price.)

The tendency of profitability, which is affected by changes of factors of derivative plants, i.e., those for producing LDPE, HDPE, PP, is similar to that of a whole complex as stated in 4-5-2, Part I. The effect of domestic price level on profitability is the greatest, so that a 10% decrease of the domestic price deprive the projects of their viability, even if the expansion of domestic demand, due to the lowered price, is taken into account. The estimated domestic demands due to the change of prices are shown in Figs. 6-4 through 6-6.

The next most serious effects are exerted by the construction costs, basic raw material prices, and intermediate raw material prices, all of which seem to be on an approximately the same level. The effect of the operational rate on the profitability is the smallest for the same reason described in 4-5-2, Part I, and, because of the production cost, the degree of the effect is the same level as the one for raw materials.

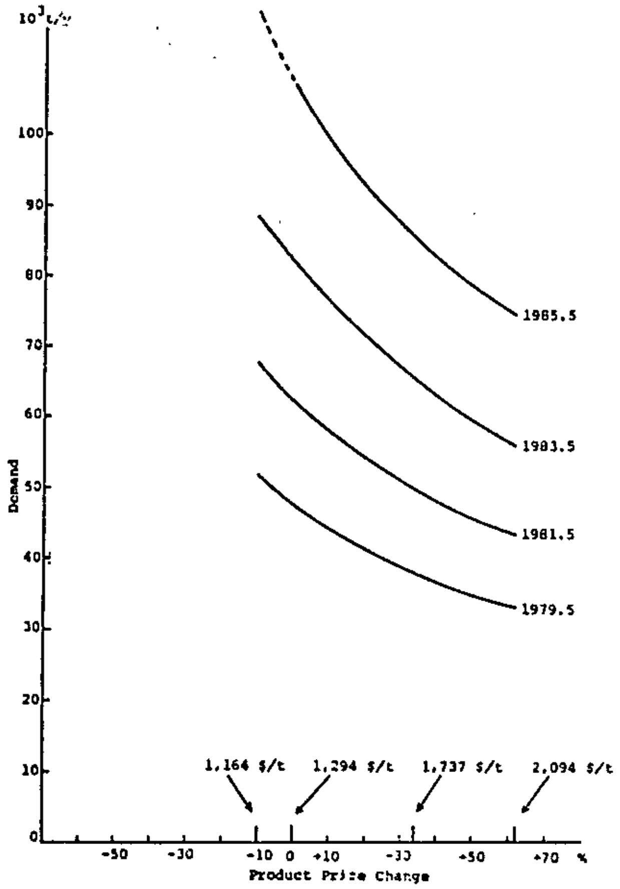


Fig. 6-4 Change of Domestic Demand of LDPE by Product Price Change

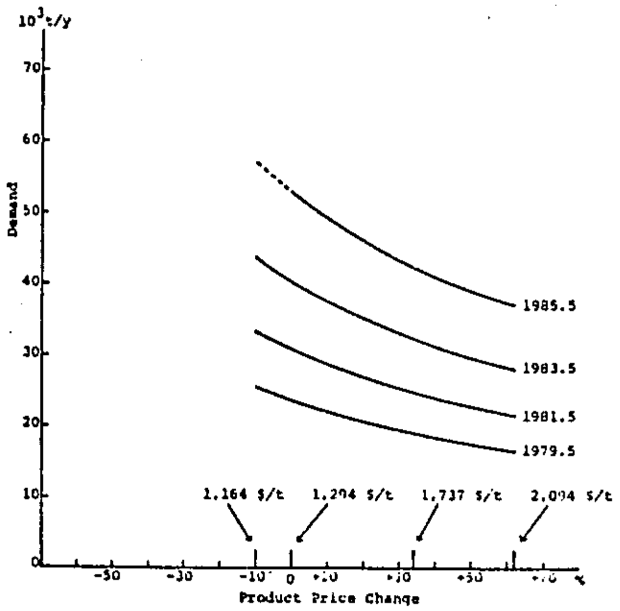


Fig. 6-5 Change of Domestic Demand of HDPE by Product Price Change

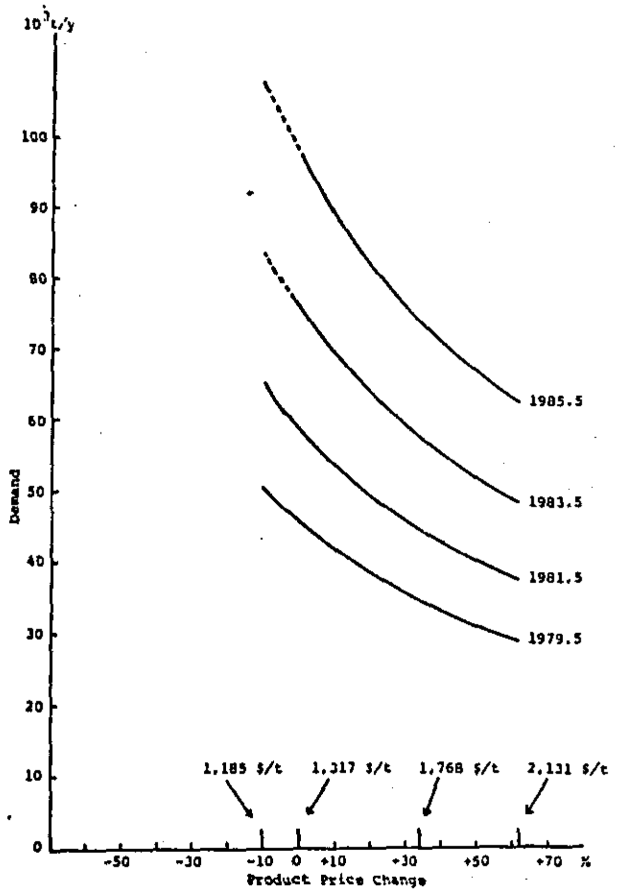


Fig. 6-6 Change of Domestic Demand of PP by Product Price Change

(1) Ethylene

As described in introduction, the sensitivity analysis is conducted integratedly. Therefore, except the case of ethylene price change, the internal rate of return is constant, but the production cost and ex-factory price of ethylene varies due to changes in the other factors (Ref. Figs. 6-7 and 6-8).

Ethylene is a basic intermediate raw material, and the price of ethylene has a serious effect on the profitability of derivative plants. Therefore, the level of ethylene price should be decided appropriately. One of the essential factors of the ethylene price is the profit rate of the ethylene plant. It is a generally accepted idea that the profitability of an intermediate plant is different from that of a derivative plant. The reason is that in comparison with the production of monomer having relatively simple ingredients, the production of derivatives, i.e., plastic resins, requires developments of applications and marketing, so that the extent of risk is higher than that of monomer production. If the internal rate of return of ethylene plant is 12%, the price of ethylene and propylene, and internal rate of return of each ethylene derivative plant are as follows:

Ethylene

Items	Basis	Possible case	Remarks
IRR	15%	12%	
Ethylene price	558 US$/t	511 US$/t	
Propylene price	446 US$/t	409 US$/t	(Polymer grade)
IRR of LDPE	16.0%	19.2%	
HDPE	13.4%	17.2%	
PP	22.5%	25.0%	
VCM	15.0%	17.1%	

As to the production cost of ethylene, raw material cost has the most serious effect, followed by operational rate. The degree of change of factors corresponding to the US$1/t price change of ethylene is as follows:

Items	Basis	Changes of factors per 1 US$/t ethylene price
Cost of raw material	151.6 US$/t	0.36 US$/t
Operational rate	90%	-0.97%
Construction cost		0.85×10^6 US$
Propylene price*	446 US$/t	2.0 US$/t

Note: * In case of changing propylene price individually

The possible deduction of ethylene price is made by (a) the modification of propylene/ethylene price ratio (ref. (4), 6-2, Part II) and (b) the conversion of relatively high price sea water to circulating cooling water (Ref. Table 4-23, Chapter 4, Part I).

(a) If the propylene/ethylene price ratio is assumed 90%, the prices of ethylene and propylene are US$538.3/t and US$484.5/t respectively.

(b) The confirmation of availability of well water has not been conducted in this study, so the necessary cooling water for ethylene plant is preliminarily assumed to be supplied by sea water system which required a greater investment. If a circulating cooling water system can be applied to ethylene plant, the following reduction of cost per ton of ethylene is possible:

451 t/t x (0.103 US$/t - 0.03 US$/t) = 32.9 US$/t

Therefore, the decreases of ethylene and propylene ex-factory prices are calculated on the allocation of reduced cost to them, US$22.6/t and US$20.4/t respectively.

Based of the above factors, the possible amount of ethylene price reduction is (557.7 - 538.3) + 22.6US$/t = 41.3 US$/t ≒ 40 US$/t

Note: Ethylene price calibration due to propylene price change is calculated in the following, as to the base case complex.

$$(557.7 - x)\,\$/t \times 203{,}000\ t/y = (-446.2 + y)\,\$/t \times 90{,}000\ t/y$$

where x, y are prices of ethylene and propylene respectively.

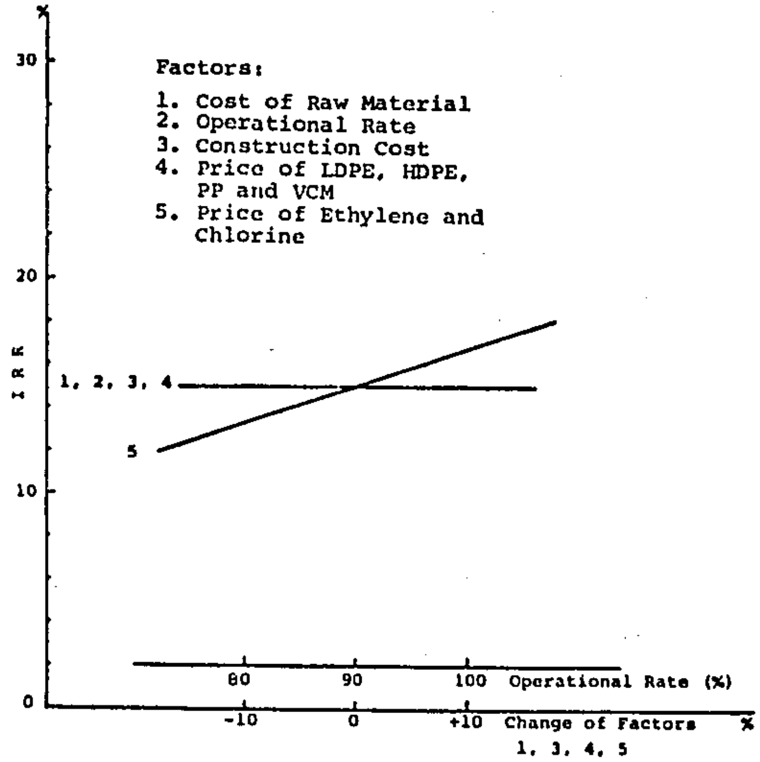


Fig. 6-7 Sensitivity of Internal Rate of Return of Ethylene to Various Factor Changes

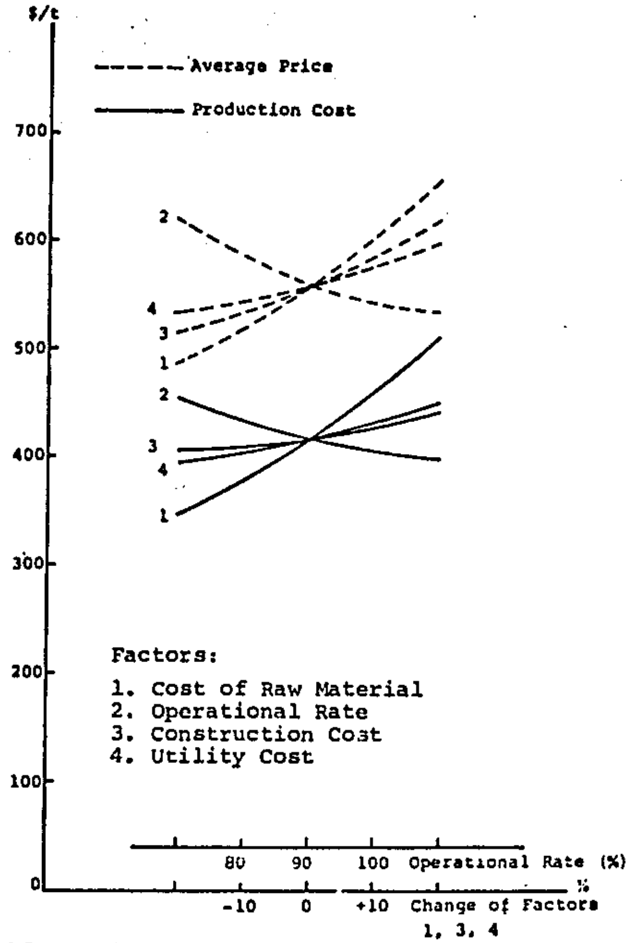


Fig. 6-8 Sensitivity of Production Cost and Average Price to Various Factor Changes - Ethylene

(2) LDPE

LDPE is the most consumable plastic resin, so that the production cost competitiveness is essential. Furthermore, as shown in Fig. 6-9 the profitability of LDPE plant is sensitive to the change of the price level, so the decrease of only 2% of the price limits the viability of the project. These situations imply the necessity of cost saving which will be accomplished by the reduction of ethylene price, construction cost, etc. The following table shows the extents of change of factors corresponding to the 1% change of internal rate of return.

LDPE

Items	Basis	Changes of Factors	
IRR	16%	1%	-
Construction Cost	-	-2.5 %	-
Ethylene price	557.7 US$/t	-2.5 %	14 US$/t
Naphtha price	151.6 US$/t	-3 %	5 US$/t
Operational rate	90 %	8 %	-

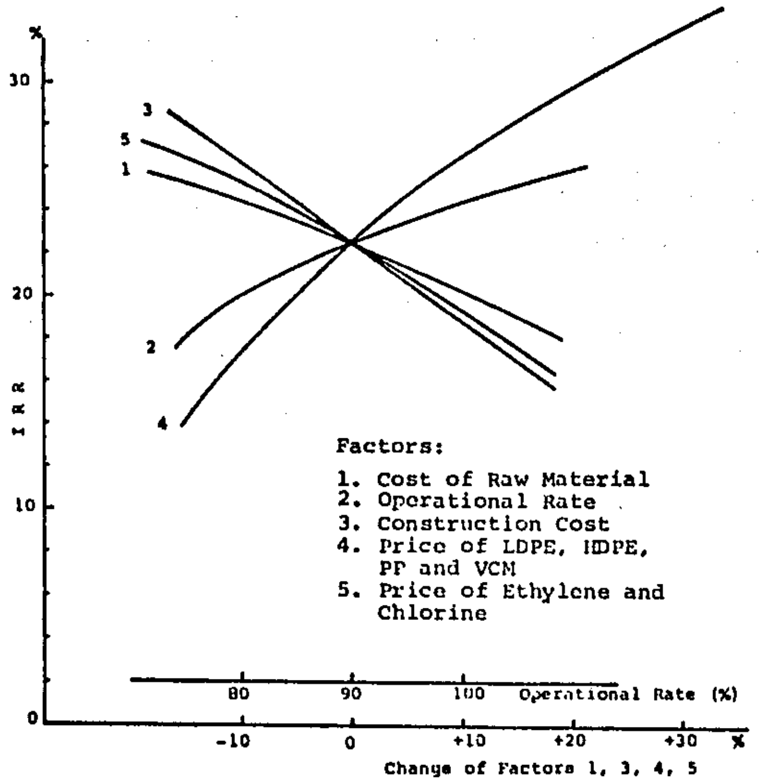


Fig. 6-9 Sensitivity of Internal Rate of Return of LDPE to Various Factor Changes

The export price of LDPE, US$681/t, is lower than the variable cost as well as production cost, so that the rise of operational rate does not afford a remarkable rise of profit. The rise of profit is mainly brought by the decrease of ethylene price only due to the rise of operational rate of the ethylene plant. The sensitivity of production cost of LDPE in Fig. 6-10 shows ethylene price and operational rates are the most critical factors.

Minimum Export Price of LDPE

Even if LDPE is exported at a lower price than the variable cost, the profitability of overall complex may increase due to the rise of operational rate in the ethylene plant. In other word, if the export price of LDPE exceeds the marginal price calculated on a whole complex basis, the profitability of the overall complex can be improved. Therefore, the minimum export price should be based on marginal price calculated on the whole complex basis as follows,

$$
\begin{array}{lcr}
m_e \times \eta_1 & = 180 \text{ US\$/t} \times 1.04 = & 187.2 \text{ US\$/t} \\
m_e & = & 139.5 \text{ US\$/t} \\
\hline
m_e^t & & 326.7 \text{ US\$/t}
\end{array}
$$

where,

m_e: Marginal price(Variable cost) of ethylene

η_1: Consumption of ethylene per product

m_e: Marginal price excl. ethylene cost of LDPE

m_e^t: A whole complex basis marginal price of LDPE

The LDPE's marginal price calculated on the whole complex basis as above is lower than the projected export price, so that the expansion of export will increase the profitability of the overall complex. However, the appropriate system for the allocation of the above profit between the ethylene and the LDPE plant should be further investigated in order to promote export.

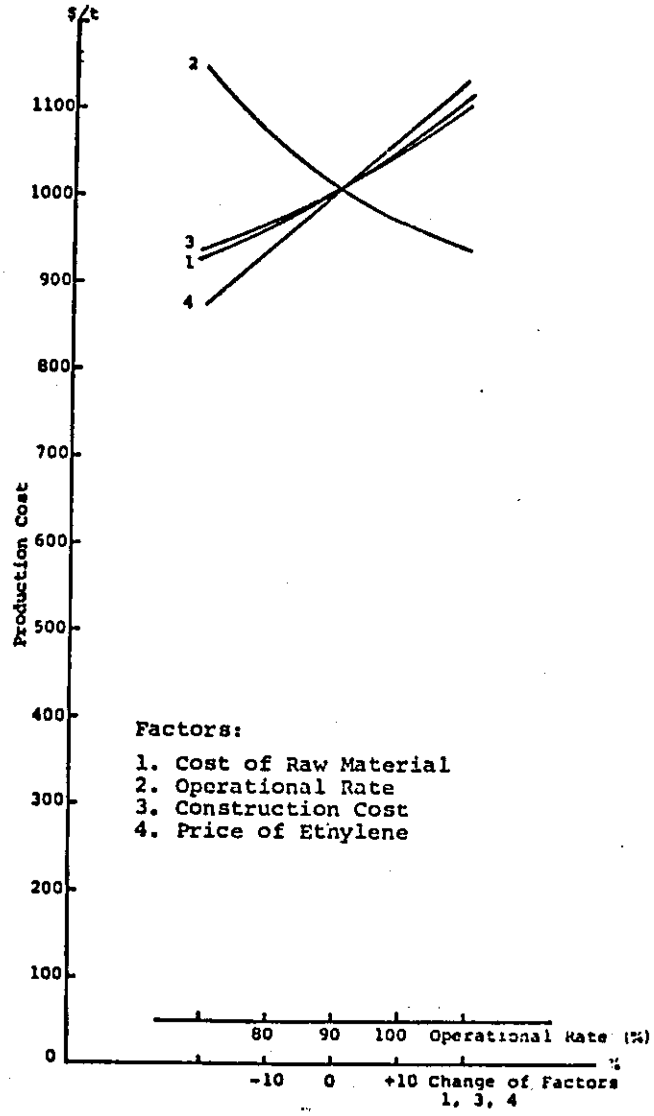


Fig. 6-10 Sensitivity of Production Cost to Various Factor Changes - LDPE

(3) HDPE

The consumption of HDPE is less than that of LDPE, so that the price is not necessarily at a low level. In comparison with the possible import price as shown in Table 4-1, Chapter 4, Part I, US$1,560/t, there exists an allowance for the rise of domestic price. The expected rise of the domestic price, US$1,395/t, required to achieve the viability of the project seems likely. The following table shows the necessary changes for the viability of the project (Ref. Figs. 6-11 and 6-12).

HDPE

Items	Basis	Charge of Factors (%)	Values of Factors @IRR = 15%
IRR	13.4%	+1.6	15%
Domestic price	1,348 US$/t	+3.5	1,395 US$/t
Ethylene price	557 US$/t	-3.8	536 US$/t
Construction cost		-4.4	-
Naphtah price	151.6 US$/t	-5.0	144 US$/t

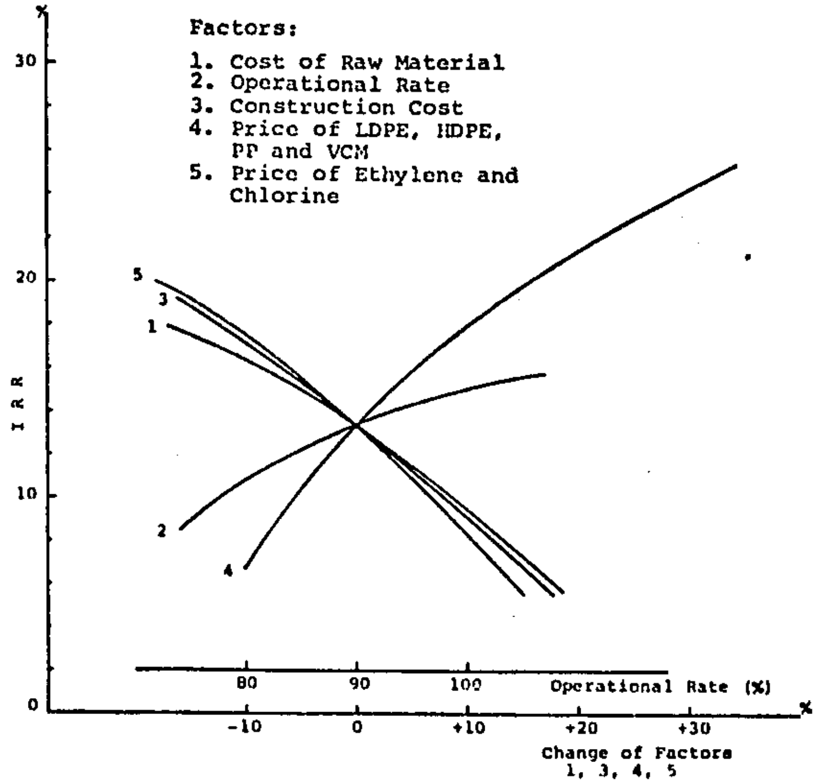


Fig. 6-11 Sensitivity of Internal Rate of Return of HDPE to Various Factor Changes

The sensitivity of production cost is shown in Fig. 6-12. The whole complex basis marginal price obtained as following is lower than the export price, US$773/t.

$$
\begin{array}{lr}
m_e \times \eta_1 = 180\text{US\$/t} \times 1.05 = & 189.0 \text{ US\$/t} \\
m_h \qquad\qquad\qquad\qquad\qquad\qquad\quad = & 199.7 \text{ US\$/t} \\
\hline
m_h^t & 388.7 \text{ US\$/t}
\end{array}
$$

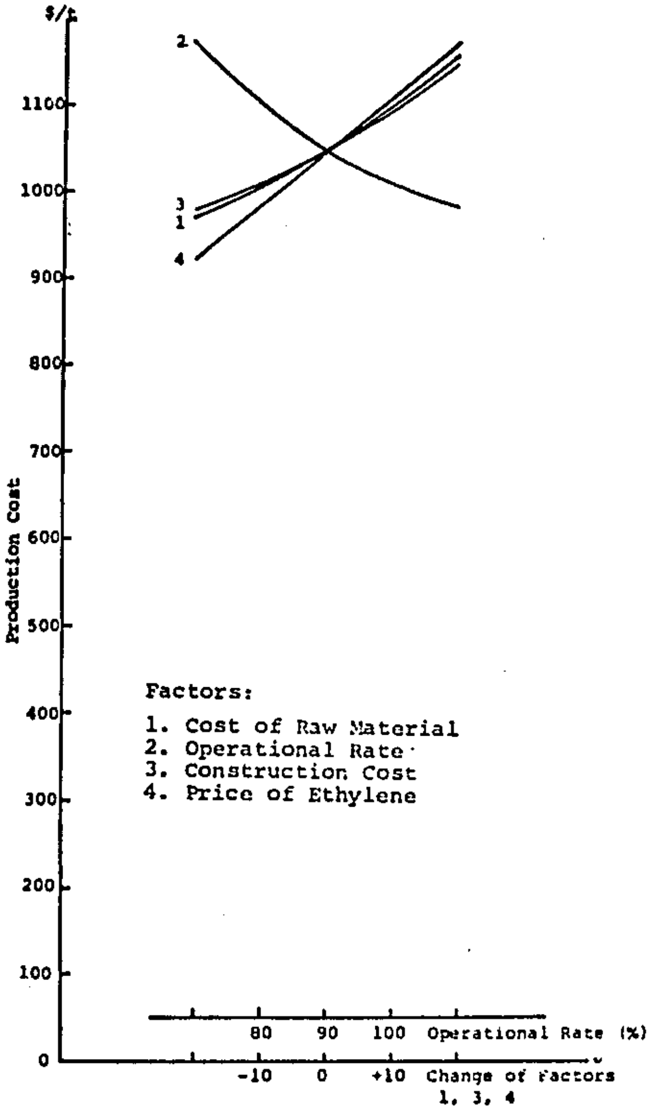


Fig. 6-12 Sensitivity of Production Cost to Various Factor Changes - HDPE

(4) PP

At present PP is the most popular plastic resin in the Philippines, so the largest demand is expected in the future. In addition to the above situation, the difference of price between propylene and ethylene may make the PP project the most profitable. However, the higher growth rate of consumption of PP in the past was recorded in the periods when the PP price is cheapest among olefinic plastic resins. On the contrary, the price of HDPE is lower than that of PP at present. Therefore, it remains a controversial question whether PP will maintain the remarkable growth of consumption in future.

The price of polymer grade polypropylene was 60 - 70% of ethylene price before the oil crisis, the difference between them decreased after the rise in oil price. In this study, 80% is settled for the propylene/ethylene price ratio as a basis of the study. Other derivatives of propylene are important petrochemical intermediates such as phenol, acrylonitrile, 2-ethyle hexanol, etc. Taking into account the possibility of producing these propylene derivatives, the decreasing of the difference between ethylene and propylene price may have to some extent limitation. In this writing, however, the allowable propylene price is studied in view of the profitability of PP project in the complex, as shown in the following (Ref. Figs. 6-13 and 6-14).

<u>PP</u>

Items	Basis	Case A	Case B
IRR	22.5%	20 %	15 %
Domestic price	1,372 US$/t	1,306 US$/t	1,185 US$/t
Propylene price (polymer grade)	446 US$/t	482 US$/t	550 US$/t
Ethylene price*	558 US$/t	540 US$/t	505 US$/t
Propylene/Ethylene price ratio	0.8	0.89	1.09

Note: Calibrated ethylene price corresponding to propylene price (Ref. Note of (1), 6-2, Part II)

As shown in the above table, it is evident that the advantageous profitability of the PP project is incurred by not the difference of price between ethylene and propylene, but by the predicted, positive market conditions. On the basis of the comprehensive consideration on the above problems, the maximum price ratio of propylene to ethylene seems to be around 0.9. It is desirable that the difference of profitability between PP project and other projects should be filled up by the effort of developing markets. The whole complex basis marginal price is obtained as follows:

$$m_p \times \eta \doteq (180 \times 0.8) \times 1.14 = 164.2 \text{ US\$/t}$$

$$\underline{m_{pp} \quad = \quad 188.4 \text{ US\$/t}}$$

$$m^t_{pp} \quad 352.6 \text{ US\$/t}$$

(Ref. (2), 6-2, Part II)

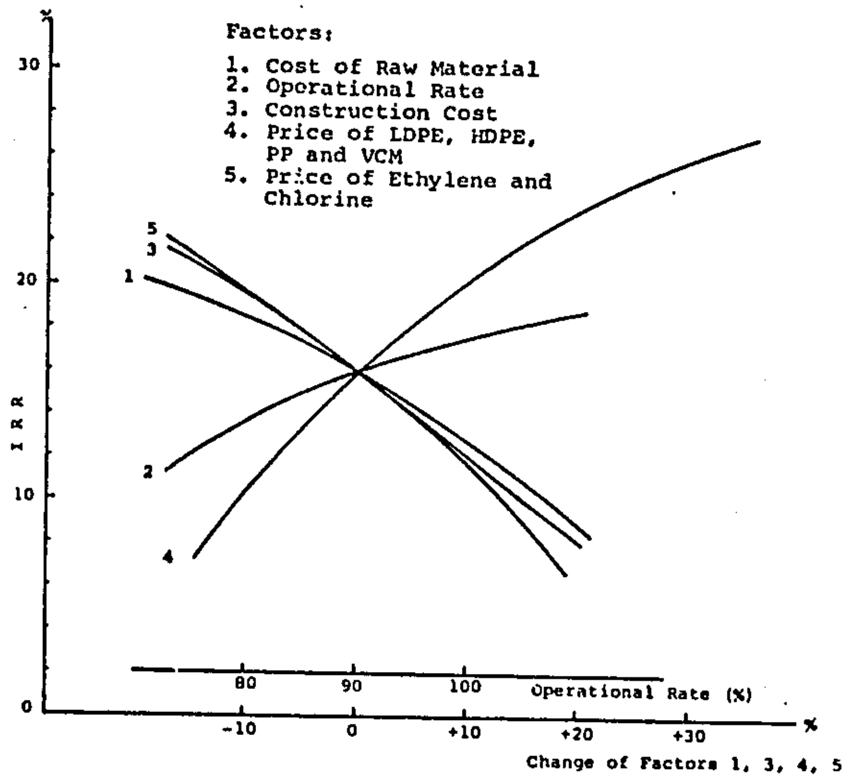


Fig. 6-13 Sensitivity of Internal Rate of Return of PP to Various Factor Changes

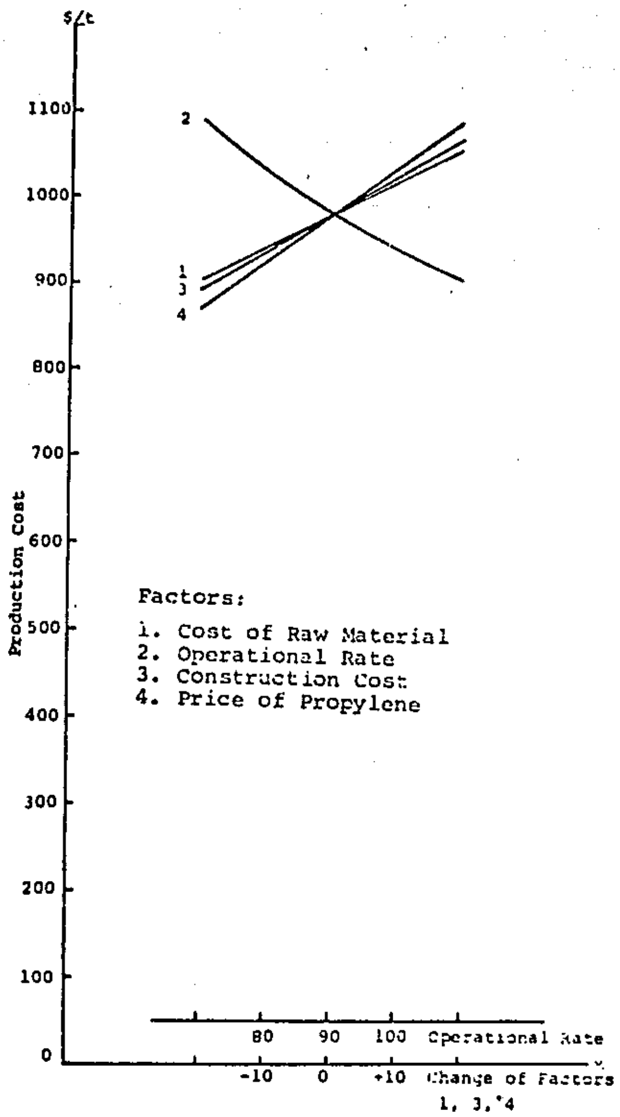


Fig. 6-14 Sensitivity of Production Cost to Various Factor Changes - PP

(5) Chlorine

Nowadays, the production process of salt electrolysis is changing from a mercury cell process to a diaphragm cell or an ion exchange membrane process. Since there is grave uncertainty regarding the essential factors such as the pollution problems, the quality of by-produced caustic soda, etc., the precise evaluation of process economics is difficult at the present. Furthermore, the prices of caustic soda and electricity depend on the specific local situation of the Philippines, so that they might be variable according to the change of situation.

In this study, the diaphragm cell process is pressumed as an electrolysis process and the by-produced low purity caustic soda is assumed to be evaporated to enrich and rectified to obtain equivalent quality to that of the mercury process.

At the present, the ion exchange membrane cell is being developed among many companies, and Asahi Chemical Ind. Co., has been operating the first commercial plant since April in 1975, in Japan. The ion exchange membrane cell process is expected to reduce the construction and production cost with the better or equivalent quality of products to mercury process.

In comparison with the imported VCM, the domestically produced VCM is not competitive as described in (6) "VCM". Therefore, for the reduction of VCM cost, the possibility of reducing the price of chlorine is studied in this paper. As illustrated in Figs. 6-15 and 6-16, the possbile changes of factors and the corresponding changes of chlorine, VCM price is shown in the following:

		Chlorine		
Items	Basis	Possibe Value	Change of Cl_2 Price	Change of VCM Price
IRR	15%	12%	-44	-26
Caustic soda price	400 US$/t	500 US$/t	-120	-72
Electricity	6.7 ¢/KWH	3 ¢/KWH	-125	-75
Construction cost		-30%	-98	-58

For the establishment of competitiveness of domestic VCM to the imported VCM, the required decrease of chlorine price is apporximately US$150/t, i.e., chlorine price should be reduced from US$256/t to US$106/t. Judging from the above table, the reduction of chlorine price to the desired level may be possible.

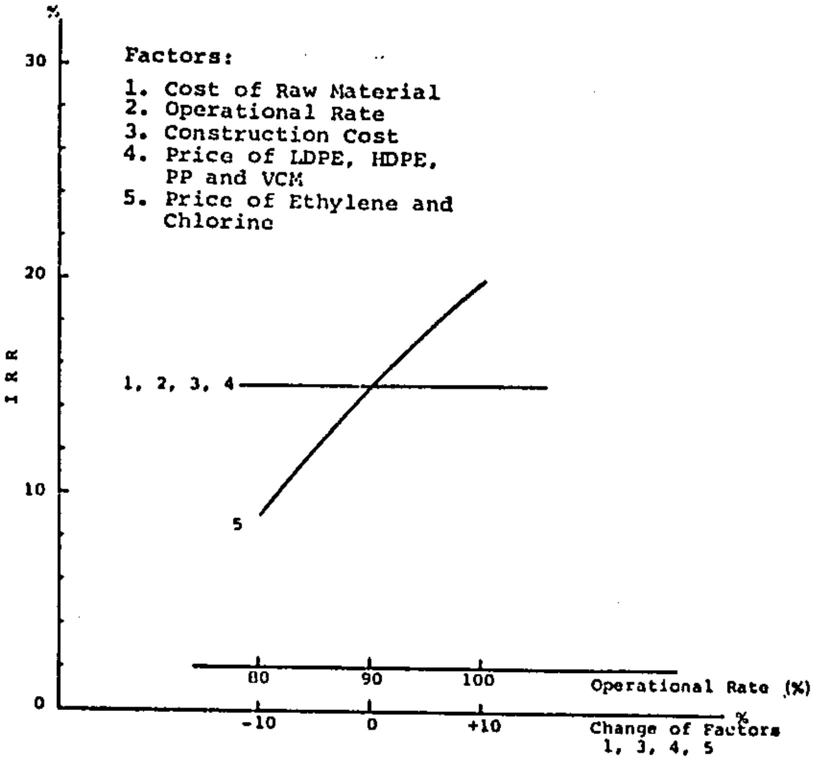


Fig. 6-15 Sensitivity of Internal Rate of Return of Electrolysis to Various Factor Changes

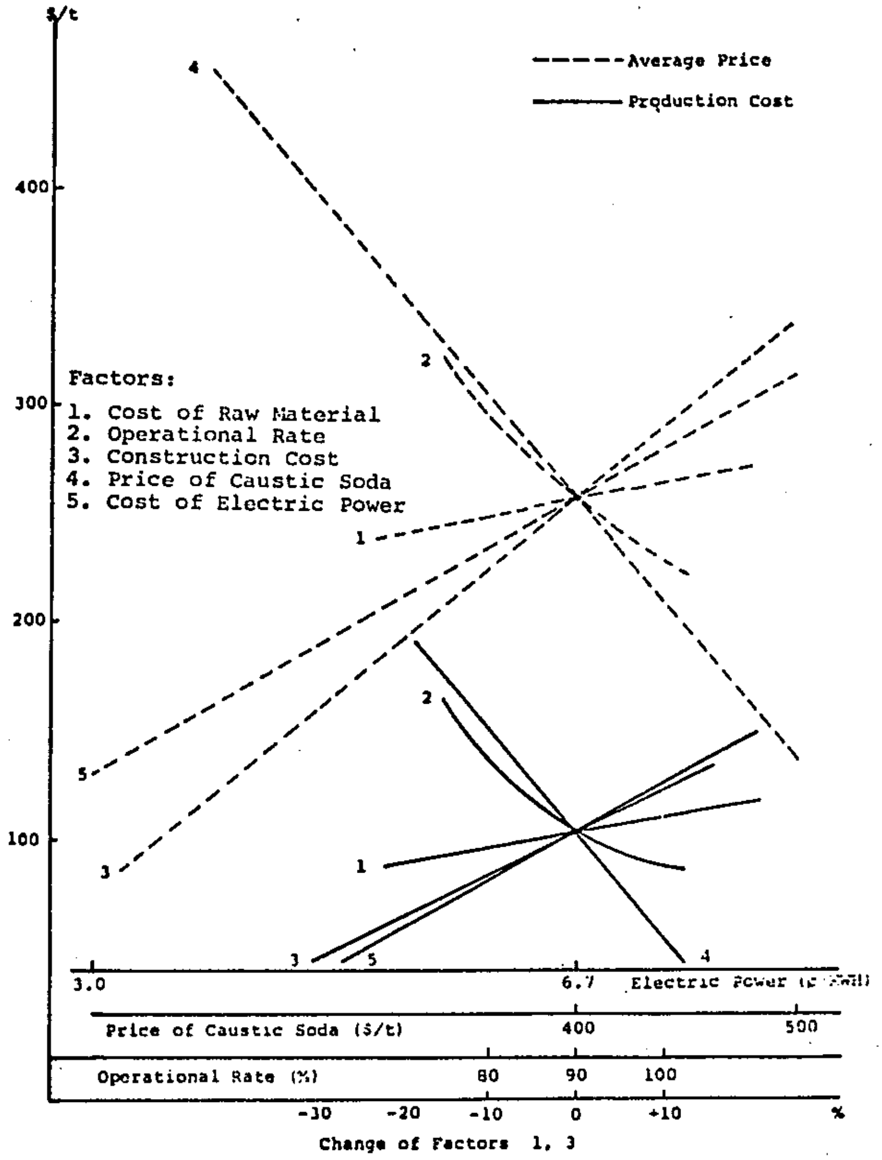


Fig. 6-16 Sensitivity of Production Cost and Average Price to Various Factor Changes - Chlorine

(6) VCM

Toxicity of VCM effecting the working of manufacturing plant has attracted the attention of authorities who are beginning to control the concentration of VCM in the environment of operating unit. Also, industrial waste, including chlorine compound, causes difficulties in 2nd stage or 3rd stage disposal treatment in public, so that the waste should be treated in the closed system.

The extent of the control of toxitic environment or the regulation of waste disposal depend to large extent on the specific local conditions of the considering country. However, in this paper, the plant cost is estimated including all of the facilities necessary for pollution control.

Domestically produced VCM calculated in the base case is expensive in comparison with the imported VCM, due to the high price of ethylene, relatively small size of plant, etc. Therefore, the possibility of VCM cost down is studied on the basis of Figs. 6-17 and 6-18. The changes of VCM price in ratio to the possible changes of other factors are shown in the following table.

VCM

Items	Basis	Possible Values	Changes of VCM price
IRR	15 %	12 %	-30 US$/t
Ethylene price	557.7 US$/t	517.7 US$/t	-20 US$/t
Chlorine price	255.7 US$/t	105.7 US$/t	-93 US$/t
Construction cost		-30 %	-66 US$/t

The price of VCM can be lowered by approximately US$180/t if the reduction of ethylene and chlorine price, construction cost are realized, and the VCM price, US$568/t = US$748/t - US$180/t, is competitive against the imported VCM.

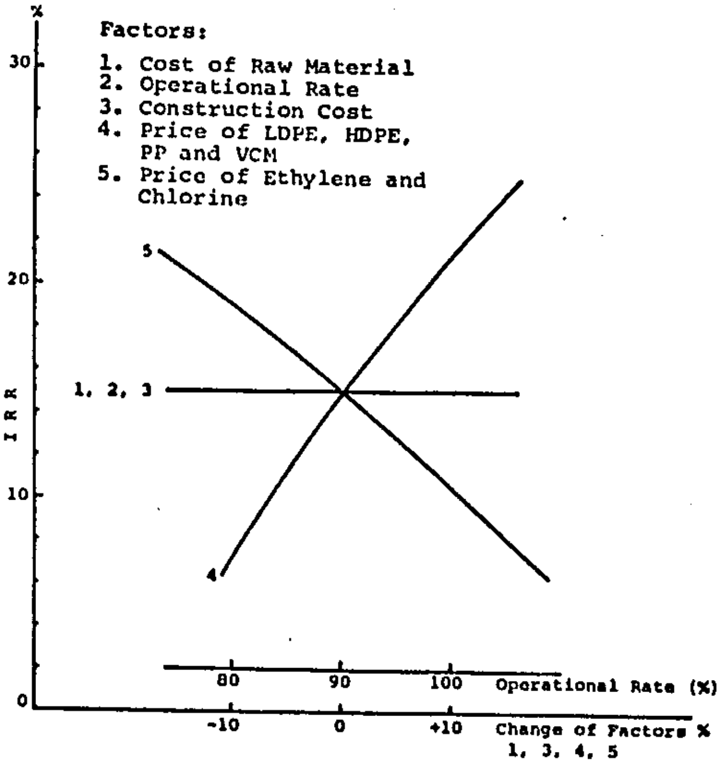


Fig. 6-17 Sensitivity of Internal Rate of Return of VCM to Various Factor Changes

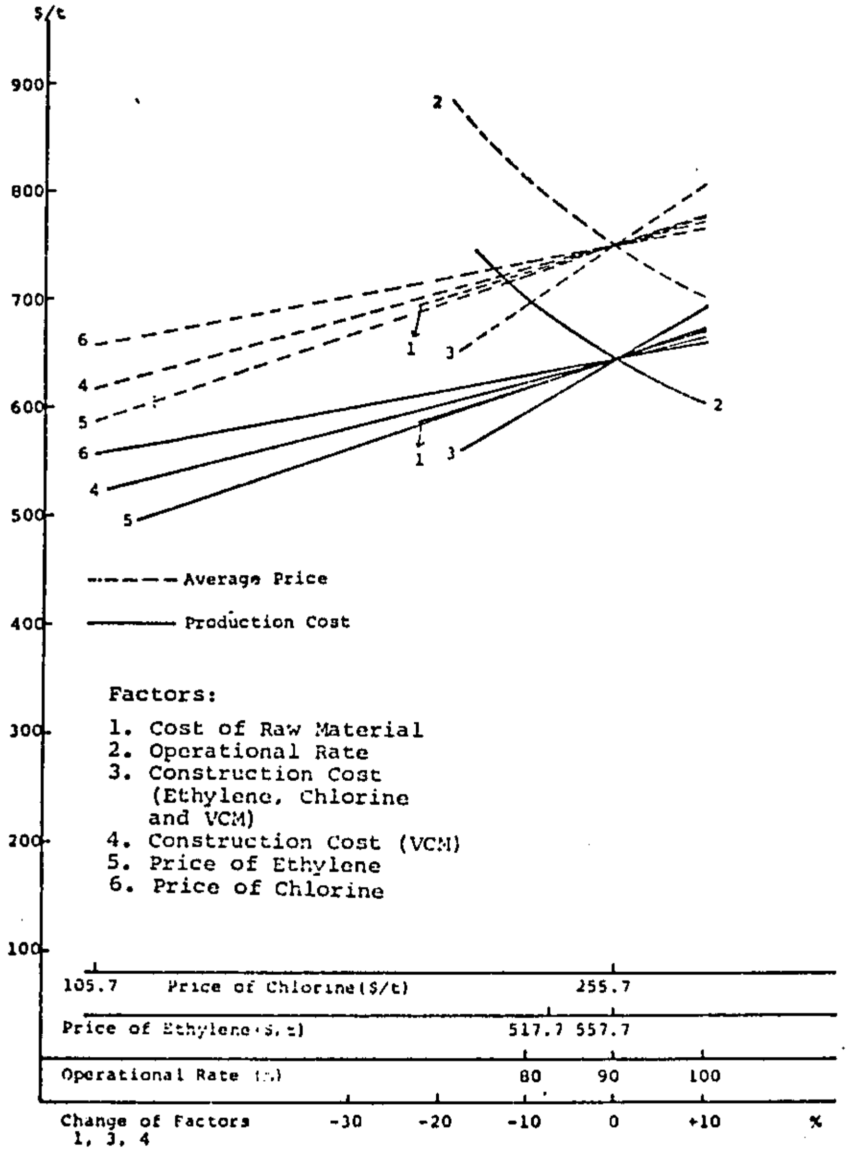


Fig. 6-18 Sensitivity of Production Cost and Average Price to Various Factor Changes - VCM

CHAPTER 7. SUMMARY OF ORIENTATION STUDY

7. Summary of Orientation Study

7-1 Introduction

(1) This Orientation Study Report has been prepared by the Consultants by appointment by the Japan International Cooperation Agency in accordance with an agreement made between the Japanese Government and the Philippine Government. This report has been compiled as part of STUDY ON DEVELOPMENT OF PETROCHEMICAL INDUSTRIES IN REPUBLIC OF THE PHILIPPINES.

The studies by the Consultants started on 1st February, 1975, and field survey mission composed of eight (8) members headed by Mr. T. Chino (UNICO International Corporation, Tokyo) was sent to the Philippines.

During the field survey in the Philippines, a back-stopping group of the Consultants in Japan prepared this Orientation Study on the basis of assumed conditions.

(2) In accordance with the major objectives of this Orientation Study which is to clarify the global prospects on the establishment of a petrochemical complex in the Philippines, the widest possible range of relative factors and alternatives have been taken into account. Further, in view of future prospects of the related chemical industries which operate on similar petroleum feeds, the establishment of an olefin complex is not the only subject of this successive studies. An aromatics complex and/or synthetic fiber raw materials complex, as well as a nitrogenous fertilizer plant, have also been studied in certain depth in this report.

In this connection, a preliminary study on a plan for allocation of hydrocarbon feeds to the related industries has been made on the basis of Filoil Refinery in Cavite which is tentatively nominated as a candidate project site.

(3) Therefore, in this Orientation Study, major subjects of study are Olefin, Aromatics and Synthetic Fiber Raw Materials, Nitrogenous Fertilizers and Allocation of Hydrocarbons. However, in regard to the related industries (i.e., aromatics, synthetic fiber raw materials and nitrogenous fertilizers), comments and recommendations by the Consultants on the results of the Orientation Study are provided herein for further elaborated studies to be conducted separately.

(4) Due to constraints of the allowed amount of time, assumed conditions employed in this Orientation Study may not be fully appropriate for this particular project study. However, in further extended studies to be conducted later, conditions specified in the "BASIC CONDITIONS FOR STUDY" attached hereto will be used. These basic conditions shall be established through mutual discussions between the parties concerned on the parts of the Philippine Government and the survey team sent by JICA to the Philippines. Therefore, it should be understood that conceptual figures and/or comparative results brought about by this Orientation Study are the essential factors for the further studies.

7-2 Summary and Comments

7-2-1 Feedstock for Petrochemical Industries

(1) Allocation of local hydrocarbon raw materials

(a) Summary

a) The expected total amount of naphtha, kerosene, and light gas oil available from the expanded Filoil Refinery (50,000 BPSD) is sufficient for covering approximately 260,000 t/y of ethylene production.

b) Available naphtha for the aromatic complex from the expanded Filoil Refinery is sufficient for approximately 40,000 t/y of p-TPA production which will be still below the expected local demand in the year 1980 (i.e., approx. 50,000 t/y as p-TPA).

c) Any fraction except LPG from the expanded Filoil Refinery is available in sufficient amounts as the feed for the 1,000 t/d of ammonia production.

(b) Comments

a) In view of the expected scope of the domestic market and the production scale of both olefins and aromatics, it seems recommendable that the supply of the feeds from the expanded Filoil Refinery be allocated with preference to olefins production.

b) In the case of establishing both complexes for olefin and aromatics production, a naphtha shortage will have to be covered by imported naphtha or by locally produced naphtha to be supplied by other refineries.

c) In order to enable the utilization of light fractions of the raw materials for olefin and/or aromatics complex, production of ammonia and urea should be planned on the basis of fuel oil as a feed which will be made available in a large surplus.

(2) Raw material price

(a) Summary

a) The project may not be viable economically if the import duty and tax in force were to be imposed on raw materials, as the raw material cost vitally affects the production cost.

b) Specifically prepared pricing systems for feeds for petrochemical industries, as practiced in all the advanced countries, have not been established as yet in the Philippines.

c) The pricing policies adopted in the advanced countries are mostly practiced by previlege on tax and duties on such a feed.

d) The international price level of naphtha as a feed for petrochemical industries is approximately US$80 to US$90/kl.

(b) Comments

a) The prices of raw materials for petrochemical industries should be properly established in the Philippines at a level similar to that prevailing in advanced countries.

b) Import duty on crude oil and specific tax on refinery products should, therefore, be adjusted by exemption or by credit for the respective proportions to be fed into petrochemical industries, so as to realize an appropriate price level of such feeds.

7-2-2 Olefins

(1) Summary

(a) Demand forecast and prospect of domestic market

Although it is not as serious as in the other countries in Southeast Asia, the Philippines market consumption of olefin resins fell sharply due to the price increase of olefin products caused by oil crisis in the late 1973. The demand forecast made by BOI before the oil crisis must, therefore, be re-examined. The results of a preliminary examination by means of the price and GDP elasticity method indicated that, in order to maintain an economic scale of olefin complex at a high operational rate the price level of olefins should be kept at level lower than the import price level at present.

(b) International tendency of petrochemical products demand/supply

The plastic resin demands in 1974 are generally estimated to be 60% to 70% of 1973 level in Asian countries. The price of plastic resins in the Asian region; however, stopped the steep increase at the beginning of 1974. The demand is forecast to recover to 80% of the 1973 level. Consequently, it is expected that the present production capacity will amply cover the demand increase at least until 1977.

(c) Future international price prospect

In view of the ability in supply, Japan will still remain as the major exporter to Southeast Asian countries. In view of the past records of export prices from Japan, the level of FOB prices will fluctuate between the domestic ex-factory price and the marginal cost. Under the circumstances in which domestic ex-factory prices and the marginal cost increase along with the increment in the prices of naphtha and ethylene, the FOB price will not drastically decrease.

(d) Production cost for ethylene

a) The fixed cost portion in the total production cost is rather small when compared with the variable cost. Further, in evaluation of the variable cost elements, the raw material price is observed as the most essential element.

b) In the comparison of different raw materials (i.e., naphtha, gas oil, LPG and LNG), naphtha and gas oil indicated satisfactory results. LPG indicated disadvantageous results due to its high cost, and heavy fraction in LNG was found impractical due to its difficulty in economical utilization of residual proportion in large lots in the Philippines. Thus, it is suggested that both LPG and LNG be discarded in the further studies.

c) The cost difference by the production scale in the ethylene plant is found to be less significant.

(e) Production cost for LDPE

a) Although the production economy of LDPE is largely dependant on the price of primary raw materials (i.e., naphtha, etc.), the production cost of LDPE in the Philippines will be still competitive with imported LDPE which is taxed by a 30% import duty and other impositions.

b) The cost difference by the production scale in the LDPE plant is larger than that in the case of ethylene production.

(f) Operational rate and production cost

The cost difference by operational rate has an overall significance. In particular, the difference in LDPE production cost is much more significant than that in the ethylene production cost.

(g) International competitiveness of olefins production cost

a) In general, the production cost of olefins is advantageous in the oil and/or natural gas producing countries.

b) The production cost by newly established plant in Japan is higher than that in the oil/gas producing countries. However, the average production cost of Japan including existing plants will still be competitive.

c) The production cost in the Philippines will not be competitive with the production cost in the other exporting countries and that in the oil/gas producing countries.

(2) Comments

(a) In view of possible oversupply and severe price competition among the exporting countries, the domestic market projection should be made on a conservative side, without involving substantial amounts of export of olefin resins which would affect project economy.

(b) In order to realize economical operation of the complex, and to ensure contribution to the national benefit enhancement, a positive policy on market development and price control on raw materials and products will be required.

(c) Project size should be determined by considering as the essential elements higher operational rates for respective plants in the complex and scale enlargement for the downstream plants as much as possible.

7-2-3 Aromatics and Synthetic Fiber Raw Materials

(1) Summary

(a) The forecast domestic demand of synthetic fibers in the Philippines are as follows:

	1979	1980
Polyester	43,300 (ton)	47,400 (ton)
Nylon	17,200	18,200

(b) The production economics has been studied for the following cases:

a) Production of synthetic fiber raw materials based on imported intermediates, i.e., cyclohexane and p-xylene.

b) Production of aromatics and synthetic fiber raw materials based on domestic raw materials, i.e., naphtha and pyrolysis gasoline.

The production scale of caprolactam and p-TPA/DMT is tentatively established as 30,000 MTA and 70,000 MTA (as p-TPA) respectively based on the demand projection in the Philippines.

(c) Production economics based on imported intermediates indicates generally preferable figures, and in particular, production economics of p-TPA based on imported p-xylene indicates the most preferable results among the cases.

(d) Production economics based on the domestic naphtha (i.e., cyclohexane--caprolactam and p-xylene--p-TPA) indicates that, if naphtha price is given as calculated including tax and duties, the caprolactam price and p-TPA/DMT price will not be competitive with imported products.

(2) Comments

(a) The foremost priority should be given to the production of p-TPA/DMT, among other synthetic fiber raw materials, for turning out polyester fiber in the early 1980's by taking into account the growth of the domestic demand, and the potential for the development of an export-oriented textile industry.

(b) The production of caprolactam for nylon is not considered to be a preferable project for the early 1980's because of the insignificant domestic demand even then.

(c) The lowest priority should be given to the production of p-xylene, a raw material for p-TPA/DMT, based on a reformate from naphtha. This is because of the possible rise in the production cost due to the higher price of raw materials and scale of the production.

(d) In view of international competition, the production of p-xylene at a scale of 50,000 MTA is too small when compared with the announced plants for 200,000 MTA to 330,000 MTA in Taiwan and Indonesia. Further, if the p-xylene production is planned in Singapore, the plant capacity will most likely be on the same level as that of Taiwan and Indonesia. It may therefore be a better policy to delay the commencement of aromatics production, if the availability of p-xylene from foreign country is confirmed.

(3) Recommendation

It is recommended that the other subsequent studies for further elaboration including the following items be conducted.

(a) Elaboration of the market study on the synthetic fibers and their raw materials, and other aromatic derivatives, especially taking into account the development of the textile production activities into an export-oriented industry.

(b) Study on the development of production of synthetic fiber raw materials, i.e., p-TPA/DMT and caprolactam, and aromatics, i.e., B.T.X, cyclohexane, p-xylene, and o-xylene based on the market study above-mentioned.

7-2-4 Nitrogenous Fertilizer

(1) Summary

(a) Agricultural productivities of major food crops in the Philippines are extremely low, particularly in the case of rice production. Therefore, positive and long-run implementation of the Masagana 99 Program including extension of fertilization is highly desired.

(b) Conservative estimates on the nitrogenous fertilizer consumption in the Philippines in the year 1980 is given as 279,000 tons on the N-nutrient basis. It seems possible to establish a large-scale nitrogenous fertilizer plant in the Philippines.

(c) It is forecast that nitrogenous fertilizers in the international market will turn to oversupply from the last half of the decade of the 1970s and the first half of the 1980s. The presently prevailing abnormally high level of fertilizer price will gradually fall from the present point towards future.

(d) The forecast future FOB Japan price of urea is as follows, while assuming a 7% annual inflation factor.

1976	US$330/t
1980	US$200/t
1982	US$220/t

(e) The raw material-wise urea production cost in 1980 with a 10% ROI based on the raw material price without miscellaneous taxation factors can be stated as follows:

Raw material	Price ($/t)	Urea price ($/t)
Imported naphtha	160	224
Domestic naphtha	217	260
Imported LPG	270	285
Domestic fuel oil (H.S.)	131	249
Imported crude oil	133	243

(2) Comment

In comparison with the above price, the production of fertilizer on imported naphtha and domestic fuel oil on the basis of a 100% operational rate show a promising result. Therefore, establishment of a large-scale fertilizer plant will contribute to the national benefit, although some governmental subsidies will be necessary for the industry.

(3) Recommendation

As is evident from the foregoing, it is recommended that further detailed study be made concerning the imported naphtha and domestic fuel oil as major raw materials to be fed into a large-scale fertilizer plant.

7-3 Major Premises

The followings are the major premises for the evaluation of the project in the Orientation Study. These premises shall be reconfirmed and revised, if required, after discussion based on the results of the study.

7-3-1 Time of Operation and Construction

(a) The complexes are expected to commence commercial operation in mid of 1979.

(b) A period of 3.5 years is taken for the construction of the complex, and the contracts for the construction are to be concluded by early 1976.

7-3-2 Priorities in the Philippines

The petrochemical complexes are assumed to be granted by the Philippine Government with a certificate of priority industry.

7-3-3 Operation Conditions

Plant life: 15 years
Number of operation days: 330 days
Annual production: Rated capacity x 100%

7-3-4 Price Escalation

Price escalation on the required items are assumed on the basis of an inflation factor of seven (7) per cent per year on current prices through 1980. Thus it is calculated as follows:

$$(1 + 0.07)^6 \fallingdotseq 1.5$$

7-3-5 Period of Evaluation and the Basic Year

The prices, costs, etc. employed in the production cost estimation are in expected figures for the year 1980 and/or in yearly average throughout 10 years of operation starting in mid of 1979 when the commercial operation commences.

7-3-6 Financial Categories

Own capital: 30% of the fixed capital requirement
Long-term foreign loan: 70% of the fixed capital requirement
Repayment: Equally divided ten annual installments

7-3-7 Prices of Raw Materials

Prices of raw materials are assessed on the basis of the prices equivalent to the refirmed products including import duty and specific tax.

PART III

METHODOLOGY APPLIED IN THE STUDY

PART III

METHODOLOGY APPLIED IN THE STUDY

CONTENTS

CONTENTS

TABLES AND ILLUSTRATIONS

Figure

CHAPTER 1. METHOD OF DEMAND FORECAST

1. Method of Demand Forecast

1-1 Method by Trend Analysis

Demand forecasts have usually been done by the so-called trend analysis making the time to be an explanatory variable which is therefore also called time series method. This method starts the point where the chasing of demand movement and the time in the past is made. Accordingly, this method may be defined as the application of a simple empirical rules. The model is expressed by Q for the demand and t for the time:

$$Q_t = \alpha_0 + \alpha_1 t \qquad (1)$$

and is sometimes expressed by:

$$\log Q_t = \alpha_0 + \alpha_1 t \qquad (2)$$

At an initial stage when the product has only been introduced into the market, the demand function is expressed by the form of (2), and in many cases by the form of (1) when the item is put into a stable growing period.

Another growth curve expressing the demand approach towards the saturation level is used, by taking the period for demand forecast to be sufficiently longer duration and if the saturation level of a demand is obtained by some means. Such a growth curve is also used for pointing the demand saturation point through the past data. Representative examples for the growth curve are the logistic curve and the Gompeltz curve:

Logistic curve, $\frac{1}{Q_t} = \alpha_0 + \alpha_1 A^t$ (3)

Gompeltz curve, $\log Q_t = \alpha_0 + \alpha_1 A^t$ (4)

Making forecasts by using these models should be based on the premise that the time regularity of the demand factor in the past is also kept the same in the future. Therefore, such a forecasting method is no longer suitable under the changed demand structure of the plastics since the oil crisis.

1-2 Method by Econometric Model

1-2-1 Method by Correlation Analysis with GDP

Plastics demand related with other economic parameters expressing a nation's economic situation and trend such as mining and industry production index, disposable personal income, etc. These economic indicators have, at the end, a relation with the overall economic parameter, GDP. Thus the demand could be expressed by the relation with the ultimate factor, GDP. This could be explained by a concept called 'Black Box' as given in Fig. 1-1:

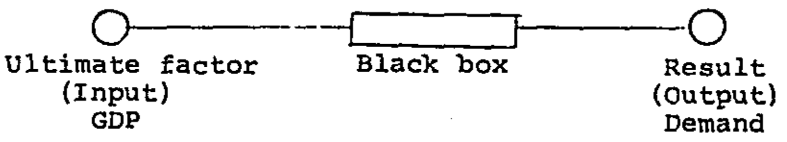


Fig. 1-1 Conceptual Figure of Black Box

The black box shows the relation between cause and effect thus providing an explanation of the output (demand) through the input (ultimate factor) macroscopically. If the structure in the black box is not changed, future predictions can be made by utilization of the relation shown by the black box.

The relationship of plastics demand, Q, and GDP Θ are expressed by the following formulae:

$$Q_t = \beta_0 + \beta_1\Theta_t \qquad (5)$$

$$\log Q_t = \beta_0 + \beta_1\Theta_t \qquad (6)$$

$$\log Q_t = \beta_0 + \beta_1 \log\Theta_t \qquad (7)$$

$$\frac{1}{Q_t} = \beta_0 + \beta_1 B^{\Theta_t} \qquad \text{(Logistic Curve)} \qquad (8)$$

$$\log Q_t = \beta_0 + \beta_1 B^{\Theta_t} \qquad \text{(Gompeltz Curve)} \qquad (9)$$

Moreover, the above formulae, (5) - (9) can be formed between per capita demand Q/W and per capita GDP Θ/W (W stands for population).

Plastics demands have been rising as the GDP rose up. However, though GDP continued rising in 1974, the plastics demands considerably diminished. Subsequently, it could not be said that this particular model is fully representative of the demand structure of the plastics.

1-2-2 Method of Multi-Correlation Analysis with an Explanatory Function of GDP and Price

Since the latter half of 1973 the plastics price have been going up turned from the then lowering trends. Plastics demands indicated a stagnant trend since 1973 due to the worldly shortage and it turned to be decreasing tendency.

Increase in the plastics demand has been supported by the GDP increase and development of related industries. However, we can add one more factor. The decreasing prices of the plastics products have replaced other materials, and expanded their market.

As the plastics price goes up, there will be a hesitance in buying plastics, and may again go back to the other materials for substitution.

Accordingly, the plastics demand will be influenced not only by the economic indicators but also by the price to great extent. This relation is given in Fig. 1-2, of which the arrow mark signifies "would give an influence":

Fig. 1-2 Demand Regulating Factors

Usually this relation is expressed by the following equation:

$$Q_t = f(\theta_t, P_t) \qquad (10)$$

It was found that the relations of plastic demand, GDP and the price in the Philippines can be expressed by the following logarithmic dual equation of the first degree as follows:

$$\log Q_t = \beta - e_p \log P_t + e_g \log \theta_t \qquad (11)$$

Where: P_t stands for real price and θ_t real GDP.
Results obtained are given in 2-1-4, Part I.

This logarithmic dual equation of the first degree is a model signifying the existence of an estimated value for the demands on the surface of 3rd degree as specifically shown in Fig. 1-3. Accordingly, if the procedure of influence coming from GDP and price against the demand does not change, the demand forecast exists on said surface or its extended surface.

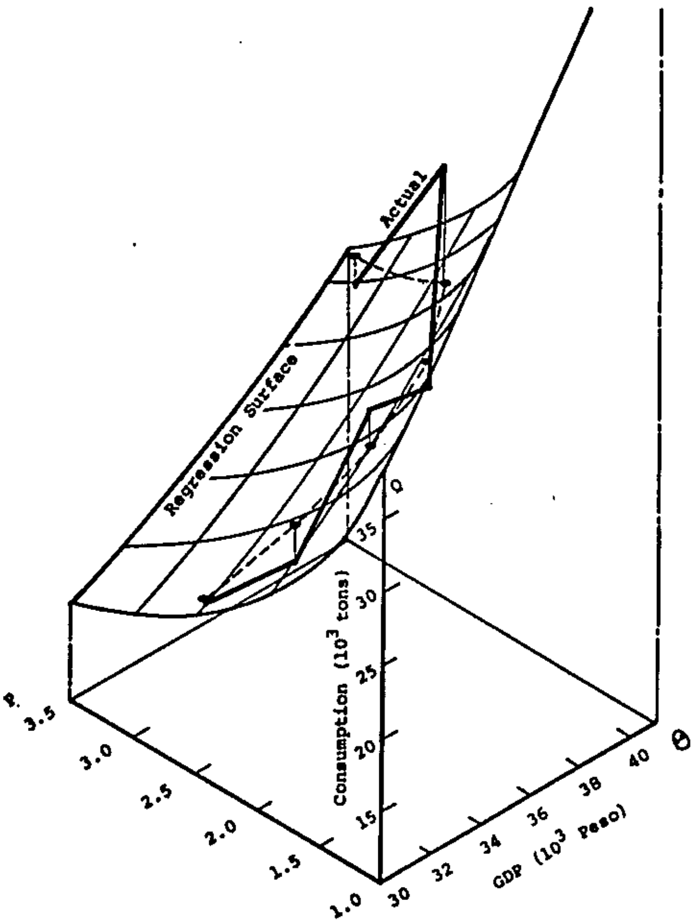


Fig. 1-3 Correlating Surface of the Demands (PE Demands of the Philippines)

1-2-3 Elasticity Analysis

The coefficients given in the above demand model (11), e_p and e_θ, are called price elasticity and GDP elasticity respectively.

The analysis of a demand structure using this kind of model is called 'elasticity analysis', and the forecast using this model generally have an assumption that the elasticity is a constant. Good matching of caluculated demands by using elasticities obtained by the analysis of the four years from 1969 to 1973 against the actual demands in the duration of 1969 to 1974 is shown earlier in 2-1-4, Part I.

However, there is no grarantee at all that this is also applicable to the coming 5 years from 1975 down to 1979. Accordingly, for the application of the model obtained through the past results for the future should make a forecast of the elasticity change towards the future.

(1) The change of the elasticity with the passage of time

Strictly this is not a change with the passage of time, but, for example, GDP elasticity changes by the size of GDP, viz.:

$$e_\theta = \hat{e}_\theta \cdot \frac{\theta_t}{\theta_t - \theta_0}$$

where: θ_0 represents GDP when the merchandise was actually introduced to the market, and θ_t the GDP in the year for which the observation was made or the year wanted to be forecast. On this occasion, $\hat{e}_\theta$ is the intrinsic elasticity and e_θ is the apparent elasticity. Similarly with e_p the following relation is obtained:

$$e_p = \hat{e}_p \cdot \frac{P_t}{P_t - P_\infty}$$

where, P_∞ stands for the lowest price of the merchandise.

(2) Forecast on the lowest demands

If the elasticity makes change in accordance with the above principle, the average elasticities of plastics in the Philippines from 1975 to 1979 reveals the followings:

Table 1-1 Projection of Elasticities for 1975/1979 (Minimum Consumption)

	GDP Elasticity		Price Elasticity	
	1969-1974	1975-1979	1969-1974	1975-1979
P E	1.89	1.17	0.68	0.59
P P	3.15	1.89	1.16	1.03
PVC	1.82	1.35	0.35	0.28

Table 2-32 of 2-1-5, Part I is the 1979 demand forecast for the year 1979 by using this elasticity.

This forecast is made with an assumption that the plastics demands in the Philippines are in the stable growing stage, and it may be regarded to be a rather pessimistic forecast. Subsequently, this is called the lowest demands.

(3) Forecast on the normal demands

As described in 1-3, Part II, the plastics demands of the Philippines are considered to be still in the rapidly accelerated developing stage, thus the demand structure is biased against particular kinds of plastics compared with that of the advanced countries. Subsequently, it may be still premature to look upon this to be the stable growth period.

Now, therefore, the value of elasticity mentioned earlier is considered to be maintainable as they are for the next 5 years.

Analyzing the Japanese plastic demand structure both the first half of the 1960s and latter half of said decade, the elasticity value during the latter half was not necessarily small. In the first half, considerably different level of elasticity prevailed depending on the kind of plastics, but this is rather turning into levelled off as it has set into the latter half.

However, in respect of PP in the Philippines, the average GDP elasticity during 1969 - 1974 has been so large that if it goes on as it is the demand becomes extremely large, which is quite questionable viewed from the olefin production balance and the by-usage demand structure. Thus assumption on the elasticity was made as below, assuming that a part of PP demands would be transferred to PE and PVC in the future.

Table 1-2 Projection of Elasticities for 1975/1979 (Normal Consumption)

	GDP Elasticity		Price Elasticity	
	1969-1974	1975-1979	1959-1974	1975-1979
P E	1.89	1.97	0.68	0.78
P P	3.15	2.20	1.16	0.93
PVC	1.82	2.00	0.35	0.81

Using the above elasticity, the normal demands are forecasted.

Fig.1-4 is the indication of the elasticity changes during the former half of 1970s and the latter half.

(4) Optimistic demand forecast

The first half of 1960s was an enjoyable period for Japan as the plastics demand increase was the most conspicuous and phenomenal of all the other period. The elasticity of every plastic material had great difference by the kind of plastic item. In average, however, the price elasticity is about 1.2 and the GNP elasticity is about 2.5.

If the real economic growth rate in the Philippines is assumed to be 7%, real price of plastic material decreases 10% during 1975, and thereafter to be lowering by 2% a year in substantiality (this assumption is identical to the occasion of lowest demand and target demand forecasts), 1.2 for price elasticity, 2.5 for GDP elasticity for each plastic material, the value as given in Part I is obtainable for the 1979 demands.

(5) Optimum demand forecast and span of forecasted value

Optimum estimation value was obtained by the following formula:

$$(\text{Optimum estimation value}) = \frac{(\text{Maximum estimation}) + 4\times(\text{target estimation}) + (\text{Minimum estimation})}{6}$$

$$(\text{Standard deviation } \sigma) = \frac{(\text{Maximum estimation} - \text{Minimum estimation})}{6}$$

Also the span of the forecasted value was expressed by:

$$\text{Optimum estimation value} \pm 2\sigma$$

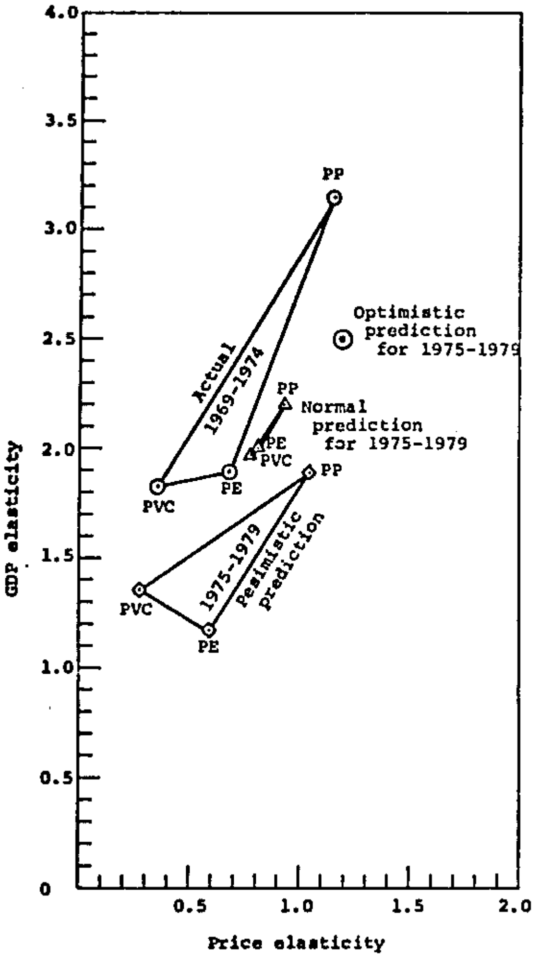


Fig. 1-4 Forecasts on Elasticity Changes

CHAPTER 2. PROJECT SELECTION

2. Project Selection

2-1 An Outline of Approach to the Evaluation and Selection of Complex Scheme through Use of a Computer Model

2-1-1 General

A computer model simulating the economic features of complex projects is employed for the evaluation and comparison of petrochemical complex alternatives. Sensitivity analysis of profitability, and production costs is also conducted by employing a computer.

Separate and detailed financial and economic analysis are made on the selected representative complex schemes.

The computer model calculates the material and utility balance, construction cost, investment cost, yearly production cost, yearly profit and loss, internal rate of return on both project (investment) and equity, and present value of the project of each process plant and also of the overall complex.

Furthermore, this model integrates the price-GDP-elasticity model for estimation of demand of petrochemical products. Therefore, the model simulates the economics of complex project under the different economic environmental conditions such as the growth rate of the national economy, rate of inflation, the pricing policy of products, etc.

2-1-2 Characteristics of Complex Computer Models

The employed models of a petrochemical complex have been built through application of a generalized process simulator (GPSOP : Generalized Chemical Process Simulation & Optimization Program), with the following characteristics (Ref. Fig. 2-1):

(1) An overall complex model is built by combined integration of each unit model simulating the material balance and economic features of a process plant.

Therefore, any configuration of the complex scheme can be optionally generated.

(2) GPSOP system is equipped with the function to utilize any optional type of unit, model, so that the employed model integrates the price-GDP-elasticity model for domestic demand projection of final products.

(3) GPSOP system can prepare mathematical equations and can automatically solve them according to the optional requirements of the users. Individual variables are specified by the user. For example, the GPSOP system can compute the appropriate level of ex-factory price which meets the requested level of internal rate of return for the project according to the specification of each variable. Conversely, the internal rate of return under the given ex-factory price can be computed depending on the specifications of variables.

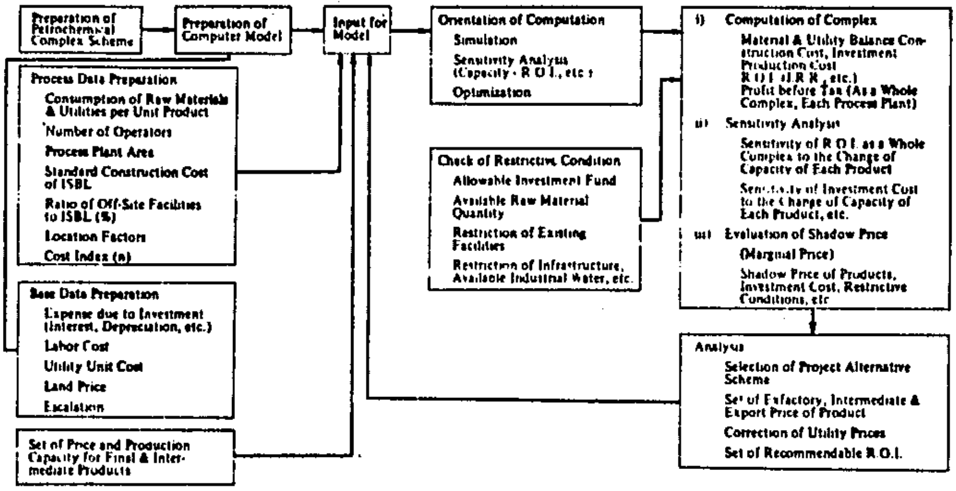


Fig. 2-1 Economic Evaluation of Petrochemical Complex by Employing GPSOP Model

2-1-3 Outline of Approach for Evaluation and Selection of Complex by Employing GPSOP Model

(1) Preparation procedures

Step (1): On the basis of the separate study on the raw materials and markets, an appropriate complex scheme is selected, as a basis, for further research of the complex scheme selection. (Ref. Fig. 2-2)

Step (2): As to the base case, material and utility balance of the complex are prepared and comprehensive studies on the construction cost, the number of operators, the land requirements, etc. of each process plant are conducted.

Step (3): Conceptual design and cost estimation of utility and common facilities of the base complex scheme are made covering the whole process plants, offices, maintenance shops and other off-site facilities.

Step (4): The common facility cost is re-allocated to the investment cost of process plants and utility plants in proportion to the number of operators, area required, and facility costs of each process plant and utility plant, according to the specific characteristics of each common facility.

Step (5): It is assumed that utilities are sold to the process plants on the basis of unit price. The utility unit price is calculated through the mutual allocation method.

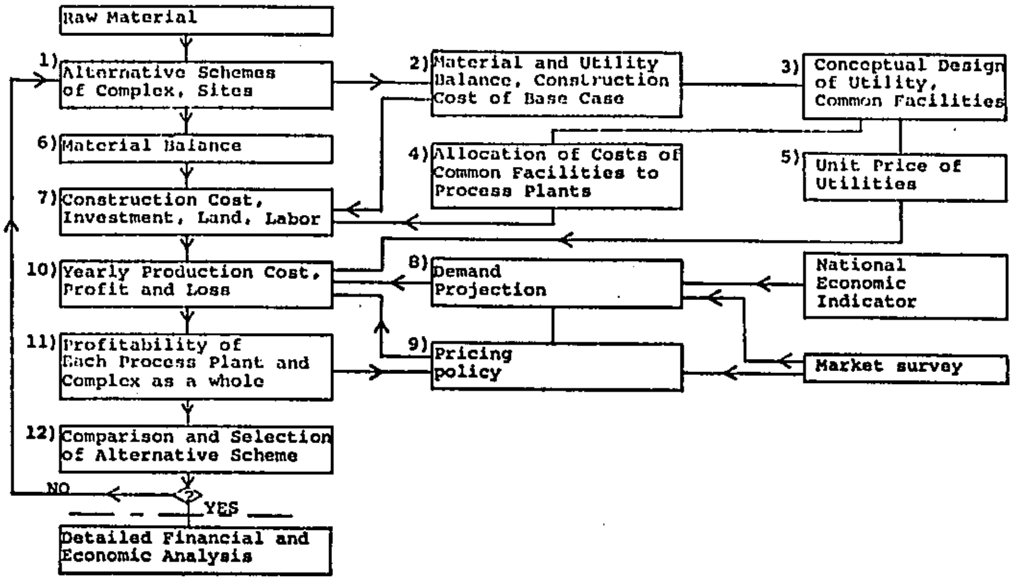


Fig. 2-2 Outline of Evaluation and Selection of Complex Schemes by Employing GPSOP Model

(2) Market projection

According to a separate market survey, the domestic demand projection is made based on the GDP (Gross Domestic Production) - price elasticity of the current domestic price and consumption of each final product. Furthermore, national economic indicators, i.e., growth rate of GDP, rate of inflation (general, each product, wage rate) are established in consultation with the proper governmental authorities.

Step (8): Demand projections for the domestic market is integratedly conducted in the GPSOP model by using the basic data and pre-requisite conditions derived from the above-mentioned studies, and the price tendencies projected by the following "pricing policy" study (step (9)).

Step (9): Pricing policy is established by taking into account both the market conditions and the profitability of each process plant. The employed GPSOP model applies the GDP-price elasticity of demand theory, so that the effects of pricing policy on the growth of demand and profitability of projects can be evaluated simultaneously during the course of economic evaluation of the complex (step (1) - step (12)).

(3) Evaluation and selection of complex schemes

Step (1): Setting up of the alternative schemes is provided for at the first stage of preparation procedures (step (2) - (5)).

Step (6): Material balance of the overall complex scheme is calculated in terms of raw materials, auxiliary raw materials, products, and by-products, to meet the requirements of the study, such as the maximum utilization of the by-products propylene.

Step (7): Construction costs, investment costs, land costs, and labor costs of each process plant are estimated by calibrating the standard data by scale factors, location factors, escalation factors, etc. The off-site facility cost of each process plant is estimated by calculating the rate against the ISBL.

Step (10): Yearly production costs, profit and loss, gross and net cash-flow etc. are calculated through the methods previously stated in 4-2-2 of this Part, to serve as a basis for economic evaluation.

Sales are calculated in the following manner:

a) Domestic demand and price is estimated on the basis of the price-GDP elasticity model by incorporating the pricing policy step (9), the growth rate of the economy, and the rate of inflation.

b) Export amount is the surplus of production minus the domestic demand. The price for export is estimated at less than the level of FOB Japan price.

Step (11): On the basis of yearly production cost and loss calculation, the profitability of each process plant and overall complex is calculated as to the internal rate of return on project (investment), equity, and the present value of the project.

2-2 Criteria for the Selection of the Economically Optimum Alternative

Generally, (a) the internal rate of return and (b) the present value of the project are applied as the criteria for the selection of an optimum project from various alternatives.

Deciding upon the criteria for the selection and development of optimum projects still remains a controversial issue among experts and specialists in the field managerial decision-making.

For example, the efficient utilization of investment is achieved through selecting a project which yields the highest internal rate of return, however, there is no guarantee that the project will yeild the maximum profit or will optimally utilize capital resources.

It is generally accepted that a project which yeilds a profit rate equal to, or greater than, the standard profit rate (cut off rate) is worth investing cost. Basically, the allocation of capital is controlled through the adjustment of the level of standard profit rate; therefore, a project that meets the standard profit rate is generally provided with the capital allocation which it requires.

On the basis of the above understanding, the creterion for the selection of economically optimum alternative has been derived as follows:

In view of the profitability, a project can be defined as a series of annual investments and returns (I^i, R^i) for i=1st through i-th year. Therefore, the difference of investments and returns ($\Delta I^i = I^i_2 - I^i_1$, $\Delta R^i = R^i_2 - R^i_1$) between an alternative project (I_2, R_2) and an existing project (I_1, R_1) can be regarded as the investments and returns of an independent single project ($\Delta I, \Delta R$), (hereinafter, the project ($\Delta I, \Delta R$) shall be referred to as a trajectory project from 1 to 2, which consists of the difference of investments and returns between a project 1 and a project 2). Therefore, if the trajectory project requires an additional investment*, the necessary condition for the feasiblity of the trajectory project ($\Delta I, \Delta R$) is the satisfaction of the following equation which is derived from the basic understanding discussed earlier.

$$r_m(\Delta I, \Delta R) = r_m(I_2 - I_1, R_2 - R_1) \geq 2s \qquad \text{................(1)}$$

Furthermore, the necessary conditions for the optimum state of the selected project, the trajectory projects from alternative projects to the optimum point should be feasible projects. In other words, the necessary conditions for the optimum state of the selected project is that the following equation (2) is effective for all alternatives except for the selected optimum:

$$r_m(I_{opt} - I, R_{opt} - R) \geq s \text{ for } (I,R) \in G, \ \&(I,R) \neq G_{opt} \text{.....(2)}$$

It is also generally observed that in actually the function in calculating the present value of the project as shown in equation (3) is a univalue, monotonic decreasing function as to the discount rate, and where an internal rate of return or a discount rate are effective as an index of profitability of a project.

$$PV(I,R,r) = \sum_{i=1}^{n} \frac{R^i - I^i}{(1+r)^i} \qquad \text{..............................(3)}$$

Therefore, the following relation is obtained:

$$r(I,R) \gtreqless r \rightleftarrows PV(I,R,r) \gtreqless 0 \qquad \text{..........................(4)}$$

On the basis of equation (1) and (4), the conditions for the feasibility of a trajectory project ($\Delta I, \Delta R$) from an alternative to optimum project is obtained as follows:

$$PV(\Delta I, \Delta R, s) \geq 0$$

$$\therefore PV(I_{opt} - I, R_{opt} - R, s) \geq 0$$

$$\therefore PV(I_{opt}, R_{opt}, s) - PV(I,R,s) \geq 0$$

$$\therefore PV(I_{opt}, R_{opt}, s) \geq PV(I,R,s) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots (5)$$

The relation (5) yields the necessary conditions for achievement of the optimum state of the selected optimum project, so that the equation (6) is effective for all the alternative projects:

$$PV(I_{opt}, R_{opt}, s) \geq \underset{(I,R) \in G}{PV(I,R,s)} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots (6)$$

It is also generally observed that inactually the function in calculating the present value of the investment (shown in equation (3)) is a univalue, monotonic decreasing function** as to the discount rate, and where an internal rate of return or a discount rate are effective as an index of profitability.

$$PV(I,R,r) = \sum_{i=1}^{n} \frac{R^{i} - I^{i}}{(1+r)^{i}} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (3)$$

Therefore, the following relation is obtained:

$$r(I,R) \gtreqless r \rightleftarrows PV(I,R,r) \gtreqless 0 \quad \ldots\ldots\ldots\ldots\ldots\ldots (4)^{***}$$

On the basis of equation (1) and (4), the conditions for the feasibility of an additional project (ΔI, ΔR) is obtained as follows:

$$PV(\Delta I, \Delta R, s) \geq 0$$

$$\therefore PV(\Delta I, \Delta R, s) = PV(I_2 - I_1, R_2 - R_1, s)$$

$$= PV(I_2, R_2, s) - PV(I_1, R_1, s) \geq 0$$

$$\therefore PV(I_2, R_2, s) \geq PV(I_1, R_1, s) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (5)$$

The relation (2) and (5) yields the necessary condition for achievement of the optimum state of the selected project, so that the equation (6) can be effective for all the alternative projects:

$$PV(I_{opt}, R_{opt}, s) \geq \underset{(I,R) \in G}{PV(I,R,s)} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots (6)$$

In other equation,

$$(I_{opt}, R_{opt}) = \left\{ (I,R) \;\middle|\; \max_{(I,R) \in G} PV(I,R,s) \right\} \quad \ldots\ldots\ldots\ldots\ldots (7)$$

Therefore, the necessary condition for the optimum state is that the selected project has to have the highest present value of a project from among all the other alternatives after being discounted by the standard profit rate.

The selected project should have the profit rate at least equal to or greater than the standard profit rate.

$$r(I_{opt}, R_{opt}) \geq s \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (8)$$

$$PV(I_{opt}, R_{opt}, s) \geq 0 \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (9)$$

The condition sufficient for the feasibility of the optimum is that the present value of the selected project discounted by the standard profit rate is at least equal to or greater than zero.

Note: (*) If the differences of investment and return between those of alternative and existing projects are less than zero, the trajectory project should be regarded as a reduction investment, on the contrary of an additional investment. In this case, it is desirable that the reduction amount of investment is greater than the reduction amount of return. Therefore, the conditions for the feasibility of a reduction investment project is :

$$r_m(\Delta I,\Delta R) \leq s \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (1)'$$

(**) monotonic increasing function for a reduction investment project.

(***) $$r(I,R) \gtreqless r \rightleftarrows PV(I,R,r) \gtreqless 0 \quad \ldots\ldots\ldots\ldots (4)'$$

On the basis of the relations between (1)' and (4)', the feasibility condition for a reduction investment project is obtained as follows:

$$PV(\Delta I,\Delta R,s) \geq 0 \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (5)'$$

Therefore, even in case of a reduction investment project, the same conclusion as an additional investment project is reached.

Nomenclature:

s	Standard Profit Rate: (Standard Rate of Return On Investment = Cut Off Rate)
r	Profit Rate: Rate of Return On Investment
r_m	Calculated Profit Rate:
I	Investment
I^i	Investment in i-th year
R	Return
R^i	Return in i-th year
G	A Set of Project Alternatives
PV	Present Value of a Project
opt	Suffix corresponding to the optimum project
1	Suffix corresponding to a certain project
2	Suffix corresponding to another project
i	Index corresponding to i-th year
n	Project Period

CHAPTER 3. CONSTRUCTION COST ESTIMATION

3. Construction Cost Estimation

3-1 Basic Consideration

(1) In general, the construction cost of industrial plants in the world have abnormally increased after the oil crisis. In particular, the prices of machinery and equipment offered by the Japanese manufactures over the past few years have increased unusually high when compared with other countries in Europe and the U.S.A. Thus, at present, the Japanese suppliers of plant equipment were no longer competitive for some time in the international market, although the price differences have finally been reduced to a negrigible range in the recently months.

(2) For the petrochemical complex in the Philippines, it is justified that the costs and prices in Japan are still applicable as the basis of construction cost estimation, because, the Japanese suppliers' prices will be competitive against European and/or American counterparts in Southeast Asia because of the geographic advantages such as lower ocean freights, quicker and reliable delivery, etc. Especially, a shorter delivery period is a vital element for shortening the construction period which is in turn essential for project economy.

(3) Therefore, the construction cost is estimated on the Japanese figures taking into account the relative conditional differences between the Philippines and Japan.

(4) The fixed-lump-sum turn-key contract with general contractor from a foreign country is assumed to be adopted as a general form of contract for the construction of each process plant (i.e., Ethylene Plant, Utility Center, LDPE Plant, HDPE Plant, Electrolysis Plant, VCM Plant and PP Plant). However, in the event of actual execution of construction, most practical and economical mode of contract should be selected in accordance with the capability of the owner.

Further, in view of the total volume of jobs required for the construction of entire complex within the pre-designated period, there may be some problem in the manpower availability in the Philippines. In general, the construction cost by application of a separate contracting system for various supplies (such as license and know-how, engineering, equipment, construction work, etc.) under management by the owner himself is lower than that of the general contracting method tentatively adopted in this study.

(5) Time of construction commencement is assumed at the beginning of 1976, and the start of commercial operation is assumed for the middle of 1979.

(6) All the construction costs being fixed at the beginning of 1976 are estimated with an escalation of seven (7) percent on the costs in 1975.

3-2 Definition and Scope of Facilities

3-2-1 Process Plants

Construction cost for each process plant is estimated by breaking down the construction cost into the following items:

(1) Equipment and materials

Machinery and equipment cost, materials for erection works (piping materials, electric wires, instrument wiring and tubing materials, other materials for construction works, etc.). Materials for structures and civil work are not included in this items.

(2) Erection work

Direct field erection cost (installation cost), i.e., all the piping work costs, machinery and equipment installation, electrical and instrumentation work, insulations, painting and coating, filling of internals, catalyst and chemicals, the mechanical-run cost up to start-up (oil-run), excluding the relative material costs. Therefore, this items shall include the direct labor, construction aid, and temporary structure costs.

(3) Civil work

Civil work cost (including the materials and construction cost) covers the buildings, foundations, structures. However, the site development or site preparation cost shall be excluded.

(4) Supervision

This item shall include all the costs necessary for carrying out supervision including the field expenses of the contractor (including indirect cost for field office, telephone, automobile rental, etc.).

(5) Engineering and contractor's fee

The engineering and contractor's fee consists of the fees required for design and engineering and the fees for general contractors. The general contractors fee includes the procurement and the inspection services, construction equipment and tools, field expenses, temporary facilities and all the necessary overheads.

(6) License and know-how

This shall include the paid up royality and basic engneering fee undertaken by the licensor, know-how fee, etc.

(7) Catalyst and chemicals

This includes the cost for solvents, coolants, and catalysts to be filled into the internals of the process equipment.

(8) Spare parts

Unless otherwise specified, the cost for spare parts for two-year operation shall be included.

The above items (1) through (5) are collectively referred to as the ISBL facilities. The cost for ISBL also covers the cost for the following items in addition to the process equipment.

Control room
Secondary electric substation
Offices in plant site
Process control analysis equipment in plant site
Receiving and handling facilities for chemicals and catalyst
Fire fighting facilities in plant site
Primary waste and waste water treating facilities
Minor maintenance shop, warehouses for special storage
Small sized receiving facilities (day tanks, etc.)

3-2-2 Utilities Facilities

The scope of the facilities which generate necessary utilities for the process plant operation is defined in 1-3, Part II.

3-2-3 Offsite facilities

The off-site facilities include process control laboratories, storage and warehouses for products and raw materials, general service facilities such as administration office, maintenance facilities, fire fighting facilities, and welfare facilities.

The scope and details of offsite facilities are shown in 1-4, and 1-6, Part II. For the estimation of the production of each product, the construction cost for general service facilities is allocated to each process plant.

3-3 Location Factors

Location factors for construction cost differ even in the same countries, if the plant locations are different. For the estimation of plant construction cost, first step is the comparison of location factors in each candidate site. However, at this stage of the study when the possibility of installing petrochemical industry in the Philippines, Rosario, Cavite is selected as a conceptional plant site.

The location factors are assumed in comparison with the construction cost elements in Japan which seem to be on an internationally reasonable cost level at present. One of the basic factors for the evaluation of location factor is labor wages and productivity. The comparison of the wage rate between Japan and the Philippines has been made through site survey by analizing labor-intensive goods production and the wage rate is assumed to be the same for both Japan and the Philippines.

3-3-1 Equipment and Materials

All the equipment and materials are assumed to be imported from foreign countries. The assumption is made on the consideration that:

(1) The major equipments are not manufactured in the Philippines in view of technical problems, and therefore importation will be absolutely necessary. The cost for such important equipment will be very high, and will occupy a large portion of the cost for equipment and materials.

(2) Some equipment and materials, the cost of which is not so high, will be manufactured in the Philippines. However, if total complex is installed during the same period, the total capacity of manufacturing equipment will not be sufficient to produce such equipment of the "Technically Manufactureable Equipment". Therefore, a considerable amount of equipment will also be imported even in this category.

(3) Cost of locally available equipment is not competitive with that manufactured in Japan, and therefore importation will not increase the plant cost..

Under the said consideration, the equipment is assumed to be imported from Japan, and the location factor is estimated to be 1.25. The factor is assessed by taking the following factors into account:

1) FOB charges and export packing in Japan
2) Ocean freight from Japan to the Philippines
3) Port handling charges for equipment and materials

The import duty for the equipment and materials is assumed to be exempted.

3-3-2 Erection Work

The cost for erection work consists of the wages (including workers' benefit) and rental fee for construction aid. Cost for the labor portion is the same as that in Japan as presumed in the foregoing. The cost difference resides in the cost for construction aid for the erection work. A considerable amount of constuction aid will have to be imported on a rental basis in view of the total availability of construction aids and requirement of handling large equipment for the complex. Further, construction period which affects the rental term is estimated to be longer than in Japan. By taking into account, the construction period, and transportation cost of construction aid, the total location factor for the erection work is estimated to be 1.2 times.

3-3-3 Civil Work

1) Cost for civil work consists of materials and labor.
2) Cost for labor is the same as the cost in Japan.
3) Total location factor is established to be 1.09 times.

Sand, gravel, cement, etc., except for steel materials, are counted as locally available items. Steel materials are assumed to be imported as the required quantity is large and the total cost is high.

It must be noted that no civil condition survey will be conducted at this stage of study, however it seems that hard layer is present one meter below the ground level as stipulated in 3-1, Part II. Therefore, no work has been taken into account, except for the foundation of the critical equipment.

3-3-4 Supervision

For the construction of plants, foreign supervisor will be required. All the local supervisors are presumed to be conducted by foreign supervisor. In addition to the total requirements of supervisors in Japan, supervisors of one for 20 laborers in erection work, and one for 100 laborers for civil work has been assumed. The cost for foreign supervisors are assumed to be US\$120 x 10^6.

3-3-5 Engineering and Contractor's Fee

Engineering fee for foreign contractor is very expensive because of uncertainty of design basis. Further, contractor's fee for foreign contractors is additionally required in this assumed basis of lump-sum contract. Average 15% of the total ISBL cost is alloted for this portion.

3-3-6 Other Items

Spare parts are estimated to be 3.5% of the ISBL cost. License and know-how fee are the same as in Japan. The same shall apply to the equipment and materials for the estimation of catalyst and chemicals.

3-4 Estimated Construction Cost

3-4-1 Process Facilities

On the basis of the said location facotrs and construction cost in Japan, the construction cost for process facilities is estimated as shown in Table 3-1. The cost for common facilities are estimated inclusively in each set of process facilities as specified in 3-4-3 herebelow.

Table 3-1 Construction Cost for Process Facilities

(Unit: 1,000US$)

		Ethylene Plant (200,000 MT/y)			LDPE Plant (110,000 MT/y)			HDPE Plant (60,000 MT/y)			VCM Plant (55,000 MT/y)			PP Plant (90,000 MT/y)			Electrolysis Plant (Cl_2 32,560 MT/y)			Total		
	Rated Capacity	Foreign Currency	Local Currency	Total	Foreign Currency	Local Currency	Total	Foreign Currency	Local Currency	Total	Foreign Currency	Local Currency	Total	Foreign Currency	Local Currency	Total	Foreign Currency	Local Currency	Total	Foreign Currency	Local Currency	Total
ISBL Facilities	Equipment & Materials	90,079	-	90,079	38,194	-	38,194	18,303	-	18,303	10,677	-	10,677	32,233	-	32,233	11,495	-	11,495	200,981	-	200,981
	Erection Works	3,103	12,410	15,513	1,079	4,315	5,394	746	2,984	3,730	429	1,715	2,144	1,403	5,610	7,013	488	1,950	2,438	7,248	28,984	36,232
	Civil Works	3,702	7,991	11,693	5,968	10,470	16,438	1,046	2,009	3,055	971	1,901	2,872	1,869	3,945	5,814	1,045	1,916	2,961	14,601	28,232	42,833
	Supervision	7,404	-	7,404	4,177	-	4,177	2,092	-	2,092	1,456	-	1,456	3,737	-	3,737	1,568	-	1,568	20,434	-	20,434
	Engineering and Contractor's Fee	22,211	-	22,211	11,339	-	11,339	4,706	-	4,706	3,074	-	3,074	8,876	-	8,876	3,309	-	3,309	53,515	-	53,515
	Total	126,499	20,401	146,900	60,757	14,785	75,542	26,893	4,993	31,886	16,607	3,616	20,223	48,118	9,555	57,673	17,905	3,866	21,771	296,779	57,216	353,995
	Spare Parts	5,143	-	5,143	2,648	-	2,648	1,114	-	1,114	886	-	886	2,019	-	2,019	762	-	762	12,572	-	12,572
	Catalyst & Chemicals	648	-	648	-	-	-	-	-	-	286	-	286	-	-	-	-	-	-	934	-	934
	Licence & Know-how Fee	1,829	-	1,829	5,676	-	5,676	4,819	-	4,819	667	-	667	7,276	-	7,276	1,724	-	1,724	21,991	-	21,991
Process Plant Total		134,119	20,401	154,520	69,081	14,785	83,866	32,826	4,993	37,819	18,446	3,616	22,062	57,413	9,555	66,968	20,391	3,866	24,257	332,276	57,216	389,492
Off-site Facilities	Process Control Laboratory	268	16	284	268	16	284	268	16	284	268	16	284	268	16	284	-	-	-	1,340	80	1,420
	Storage & Ware-house	4,489	1,858	6,347	3,067	1,234	4,301	1,591	672	2,263	453	193	646	2,389	1,012	3,401	-	684	684	11,989	5,653	17,642
	Other Common Facilities	2,949	3,195	6,144	1,968	2,133	4,101	1,472	1,595	3,067	817	886	1,703	2,107	2,283	4,390	775	841	1,616	10,090	10,931	21,021
	Total	7,706	5,069	12,775	5,303	3,383	8,686	3,331	2,283	5,614	1,538	1,095	2,633	4,764	3,311	8,075	775	1,525	2,300	23,419	16,664	40,083
Contingency Provisions		7,091	1,274	8,365	3,719	909	4,628	1,808	364	2,172	999	236	1,235	3,109	643	3,752	1,058	270	1,328	17,784	3,696	21,480
Grand Total		148,916	26,744	175,660	78,103	19,077	97,180	37,965	7,640	45,605	20,983	4,947	25,930	65,286	13,509	78,795	22,224	5,661	27,885	373,477	77,578	451,055

3-4-2 Utility Facilities

Construction cost for the utility facilities is estimated as shown in Table 3-2.

Table 3-2 Construction Cost for Utility Facilities

(1,000 US$)

Facilities	Plant			Off-site			Total		
	Foreign Portion	Local Portion	Total	Foreign Portion	Local Portion	Total	Foreign Portion	Local Portion	Grand Total
Steam Generation	14,497	4,405	18,902	502	445	947	14,999	4,850	19,849
Electric Power Generation									
Electric Generation	19,246	4,741	23,987	1,937	1,738	3,675	21,183	6,479	27,662
Sea Water Intake	6,853	8,984	15,837	5,613	1,271	6,884	12,466	10,255	22,721
Well Water Facilities	258	198	456	538	315	853	796	513	1,309
Instrument Air Facilities	1,138	491	1,629	36	33	69	1,174	524	1,698
Plant Air Facilities	476	130	606	16	15	31	492	145	637
Oxygen and Nitrogen Generation	6,992	2,665	9,657	216	201	417	7,208	2,866	10,074
Water Treatment Facilities	3,179	1,336	4,515	96	89	185	3,275	1,425	4,700
	52,639	22,950	75,589	8,954	4,107	13,061	61,593	27,057	88,650

Note: The above costs include spare parts equivalent 4,440 × 10^3 US$

3-4-3 Common facilities

Construction cost requirements for the common service facilities are shown in Table 3-3. Construction cost of each set of the common facilities is allocated in accordance with the characteristics constructions, and in proportion to the cost of ISBL, man-power, and the plant area of each process plant as shown in Table 3-1. The allocation factors are shown in Table 3-4.

Table 3-3 Construction Cost for Common Facilities

	Foreign Currency	Local Currency	Total
Administration Office	358	1,128	1,486
Fire Fighting	4,291	1,099	5,390
Parts and Chemical Warehouse	1,073	1,374	2,447
Road	-	1,620	1,620
Fence	-	116	116
Parking Area	-	101	101
Green	-	2,560	2,560
Communication	537	14	551
Road Lighting	536	14	550
Sewer System	-	174	174
Truck Scale	179	14	193
Welfare Facilities	2,503	1,200	3,703
Maintenance Facilities	5,007	1,866	6,873
Waste Water Treatment	2,682	3,138	5,820
Waste Treatment	715	43	758
Total	17,881	14,461	32,342

Table 3-4 Allocation Factors for Common Facility Cost

	Ethylene	LDPE	HDPE	VCM	PP	Electrolysis	Process Plant Total	Utility	Total Complex
ISBL %	40.6	21.0	8.9	7.4	16.1	6.0	100% (82.4)	21.3% (17.6)	121.3% (100%)
Man-Power	20.0	21.5	15.6	11.2	19.2	12.4	100% (94.4)	3.7% (3.6%)	103.7% (100%)
Waste Water	12.0	16.7	24.0	2.9	38.3	1.4	100%		
Allocation Factor									
ISBL	17.8	9.2	3.9	3.2	7.0	2.6	43.7		
Labor	3.3	3.6	2.6	1.8	3.2	2.0	16.5		
Area	3.6	3.5	3.5	3.5	3.5	3.5	21.1		
Other	2.2	3.4	4.8	0.5	7.5	0.3	18.7		
Total	26.9	19.7	14.8	9.0	21.2	8.4	100% (86.7%)	15.3% (13.3%)	115.3% (100%)

CHAPTER 4. OTHER COSTS AND PRICES ESTIMATION

4. Other Costs and Prices Estimation

4-1 Raw Material Prices

4-1-1 Estimation Basis

At present in the Philippines, the prices for petrochemical feeds are not yet established. There are many ways of determining prices of raw materials such as naphtha, gas oil, etc. They are evaluated in terms of either heating value or by subtracting, from the prices of gasoline, the cost for converting into gasoline.

However, in the Philippines, the prices of refined products are controlled by the Government as "posted prices" which are determined mainly by the policies of the Government, and not necessarily from the cost viewpoint.

When the prices are politically established, it is difficult to determine the prices of raw materials for petrochemical feed in view of production economics. Therefore, the prices of raw materials in this report will be based on the posted prices.

4-1-2 Tax and Duty System

At present, the import duty on crude oil is 20% on 110% of FOB prices regardless of the eventual utilization, and the special taxes on refined products from refineries are imposed.

However, for the imported naphtha for fertilizer plants, import duty and specific tax is deducted either wholly or partially, according to information collected by JETRO.

In the case of a petrochemical complex, the result of production cost estimation in the "Orientation Study" shows that the prices of naphtha and gas oil including specific tax and import duty on crude oil will not make the project viable.

Therefore, as a tentative basis for the pre-feasibility study, the import duty on crude oil, and the specific taxes and special fund on the refined products are assumed to be deducted when used as petrochemical raw materials.

4-1-3 Prices of Crude Oil

The price of crude oil for Kuwait Crude, which has been used as the basis for project evaluation, consists of the following cost factors in the case of Filoil Refinery. The figures have been established from the case of Bataan Refinery by modifying the ocean freight and insurance premiums.

The FOB prices of Kuwait crude is US$10.365/bbl taking 93% of posted prices of Kuwait crude.

FOB	US$10.365/bbl	
Freight LR-2	US$1.128/bbl	
Insurance 0.00225	US$0.026/bbl	
CIF price	US$11.519/bbl	
Import duty	US$2.280/bbl	(FOB x 110% x 0.2)
Wharafage	US$0.077/bbl	
OIC/loss	US$0.069/bbl	(0.006 x CIF)
	US$13.945/bbl	

The prices include import duty on crude oil. When the import duty is discounted, the prices of crude oil becomes US$11.665/bbl. This price will be used as the basis for petrochemical feeds.

4-1-4 Posted Prices and Refinery Netback

The posted prices and refinery netback are given by PNOC as of February, 1975, as shown in Table 4-1.

The gaps between the refinery netbacks and posted prices correspond to the distribution cost and storage cost at oil terminals. In the case of the proposed petrochemical complexes adjacent to the Filoil Refinery, the raw materials will be directly delivered from the refinery to the ethylene plant. Therefore, the distribution cost will be almost negligible. The prices of raw materials for petrochemical feed at the inlet of the ethylene plant will be assessed on the basis of the refinery netback after adjusting the import duty portions without taking the distribution cost into account.

Table 4-1 Prices of Petroleum Products

	Posted prices excluding tax	Refinery netbacks
LPG	0.615 Ps/L	0.0878 $/lb
Prem. gasoline	0.760 Ps/L	16.056 $/bbl
Reg. gasoline	0.705 Ps/L	14.846 $/bbl
Solvent	0.7275 Ps/L	14.601 $/bbl
Jet A-1	0.810 Ps/L	15.108 $/bbl
JP-4	0.869 Ps/L	17.125 $/bbl
Kerosene	0.720 Ps/L	15.108 $/bbl
Diesel oil	0.685 Ps/L	15.035 $/bbl
Industrial fuel oil	0.5605 Ps/L	11.828 $/bbl
Asphalt	0.6115 Ps/L	

4-1-5 Equivalent Petroleum Products to Petrochemical Feeds

In this study, the petrochemical feeds such as naphtha, kerosene, and gasoil are chosen as the subjects, and the equivalent petroleum in this respect products are as follows.

Gasoline = Naphtha

Kerosene = Kerosene

Gas Oil = Diesel Oil

Actually there is a price gap between the petrochemical feeds and the petroleum products, as it is necessary to give some treatment to produce the petroleum products from the petrochemical feed. Therefore, the petrochemical feeds can be evaluated to be less than the petroleum products. However, an estimation on the basis of the present refinery netback shows that even the prices of the existing petroleum operation are not so beneficial.

Under such a circumstance, the price of each petrochemical feed is assumed to be equal to the prices of the equivalent petroleum products.

4-1-6 Prices of Hydrocarbon Raw Materials

On the basis of the foregoing assumptions, the prices of raw materials have been estimated in the following manner:

Naphtha (gasoline equivalent):	US$14.846/bbl
Kerosene (Kerosene):	US$15.108/bbl
Gas oil (Diesel oil):	US$15.035/bbl

These prices include the import duty portion. If the import duty on these yields of products are deducted, the price of each raw material will be assessed as about 87% of each original price with the refinery cost, refinery loss, prices of fuel oil, etc. taken into account.

Therefore, a rate 87% is taken as a basis. The prices for the raw materials are as follows.

Naphtha : US$12.916/bbl = US$81.23/kl

Kerosene: US$13.144/bbl = US$82.67/kl

Gas Oil : US$13.08/bbl = US$82.26/kl

The escalation of prices is assumed to be 7% annually. The prices of feedstocks in 1980 are assessed as follows. These figures adopted as the basis for the evaluation of the Petrochemical complex.

Naphtha
(Sp.Gr. 0.724)

$$\frac{\text{US\$}81.23 \times (1.07)^5}{0.724} = \text{US\$}157.2/\text{t}$$

Kerosene
(Sp.Gr. 0.804)

$$\frac{\text{US\$}82.67 \times (1.07)^5}{0.804} = \text{US\$}144.6/\text{t}$$

Gas Oil
(Sp.Gr. 0.846)

$$\frac{\text{US\$}82.26 \times (1.07)^5}{0.846} = \text{US\$}136.7/\text{t}$$

In the case of gas oil for ethylene production, desulfurization is necessary. A typical desultory cost of US$11.0/t of gas oil is added to the prices of gas oil, thereby obtaining a price at US$147.7/t.

4-1-7 Price of Industrial Salt

The supply of the industrial salt consumed in the Philippines is met by imports from Australia and India, and by domestic supply from Pangasinan and Curanta of which the respective current prices are as follows:

Price of Raw Salt (1974)

Domestic : CIF ₱200 - 230/t or US$29 - 33/t

Imported : CIF - US$22.80/t

The expected imported salt price in 1980 is calculated as follows:

$$\text{US\$}22.80 \times (1 + 0.07)^6 = \text{US\$}34.20/\text{t}$$

Thus, the price of raw salt used in this study is assumed as US$35.-.

4-2 Evaluation of By-products

4-2-1 Basic Consideration for the Pricing of Hydrocarbon By-products

Various of hydrocarbon by-products are produced from a petrochemical complex, evaluation of which greatly affect the viability of the complex, especially that of the ethylene plant. The basis for evaluation of hydrocarbon by-products is established and follows:

(1) Hydrogen rich gas

If hydrogen can be utilized effectively for the related industry the prices of hydrogen can be evaluated at level higher than that of fuel. However, the hydrogen transportation is difficult, and the application is limited in the case of Filoil.

Therefore, hydrogen should be used as fuel within the complex or the refinery (Some part of hydrogen will be used for desulfurization of fuel oil). Therefore, the prices are estimated in terms of fuel prices.

(2) Methane rich gas

Methane is usually burned within the complex as fuel and the prices are estimated on a fuel basis.

(3) Mixed C_3 and C_4

Propane and Butane can be sold to domestic users as LPG at a price level of LPG in the Philippines. However, the domestic demand for LPG greatly fluctuates, and no constant demand is guaranteed.

Therefore, the whole amount of the mixed C_3 and C_4 are evaluated to be used as fuel within complex.

(4) Pyrolysis gasoline

The pyrolysis gasoline can be sold to other refineries as a gasoline blend by adding hydrogen because of its high octane number. This fraction is evaluated on the same level as that of naphtha.

(5) Cracked fuel oil

This fraction cannot be mixed with commercial fuel oil due to the sludge problem. This fraction can be utilized separately as fuel for petrochemical or boiler facilities.

(6) Excessive propylene

Sometimes excessive propylene will be produced due to the change in the balance of ethylene/propylene. The excessive amount of propylene is evaluated in the same manner as for mixed C_3.

4-2-2 Fuel Evaluation

The main criterion for evaluating the by-product is the fuel price. The estimation of fuel is affected by the sulfur content and heating value. As to the sulfur content, the pollution control in the Philippines at present is not stringent, and the future direction of the control strictness is difficult to forecast. Therefore, the fuel is evaluated regardless of the sulfur content, and only the heating value will be taken into account. The method for evaluating by-products is rather conservative because sulfur content in such by-products is generally smaller than those in industrial fuel oil in the Philippines.

The current refinery netback on industrial fuel oil including import duty on crude oil will be used as the basis for fuel. No deduction of import duty on this portion is assumed and the special fund and specific tax are excluded in view of the consumption within the same complex.

Conversion of industrial fuel oil price into the price per unit heating value is made as follows:

Refinery netback: US$11.828/bbl (given by PNOC)

$$11.828 \times \underset{\text{(bbl kl)}}{\frac{1}{0.159}} \times \underset{\text{(kl ton)}}{\frac{1}{0.966}} \times \underset{\text{(Heating V/ton)}}{\frac{1}{9.77}}$$

$$= \text{US\$7.81/MMKcal}$$

The figure US$7.81/MMKcal is the price in 1975 and US$10.95/MMKcal in 1980 is established by incorporating 7% annual escalation.

4-2-3 Prices of Hydrocarbon By-products

(1) Hydrogen rich gas

The guaranteed hydrogen content is 90 mol% and the expected content is higher than 95%. Therefore, 95 mol% of hydrogen and 5 mol% of methane will be the basis for evaluation.

US$261/ton

Heating methane : 11,950 Kcal/kg

Hydrogen : 28,670 Kcal/kg

Overall : 23,600 Kcal/kg

(2) Methane rich gas

The estimated composition is 94 mol% of methane and 6 mol% of hydrogen.

US$132.4/t (average heating value 12,000 Kcal/kg)

(3) Mixed C_3

US$122.2/t as fuel (Propane basis)

(4) Mixed C_4

US$120.3/t as fuel (Butane basis 10.9 Kcal/kg)

(5) Pyrolysis gasoline

US$157.2/t as naphtha

(6) Pyrolysis fuel oil

US$78.8/t as fuel (Heating value estimated 10,000 Kcal/kg)

4-2-4 Price of Caustic Soda

As shown in the information by BOI's investigation (ref. 3-4, Part III), the domestic demand for caustic soda is larger than production, and the price is much higher than in the other countries.

Example: Ex-factory Price by Domestic Producer (Philippine-1974)

US$400 - 500/t

Delivered Price (Japan - 1975)

US$150/t

The price ratio of caustic soda/chlorine (ie., 1.4 - 2.33) shows heavy dependence of costs on caustic soda due to balance of supply and demand of these materials.

Further, the price of chlorine for industrial use may be evaluated much lower.

In Japan, the price ratio of caustic soda/chlorine is approximately 1.0.

In consideration of abnormal price gap between the current market price in the Philippines and the international market price, the market price prediction on caustic soda will be made without considering any escalation and shall therefore be deemed unchanged even in 1980.

Thus, the applicable price range is presumed to be US$400 - 500/t for this study, and applied price for the financial analysis is US$417/t on an ex-factory basis.

4-3 Utilities

Investment is necessary for the generation and distribution of utilities such as steam, electric power, water, instrument air, etc. The prices of utilities are estimated on the cost basis for generation and distribution. The prices are estimated in such a manner that each set of the facilities have a 10% of Return On Investment (ROI) at an operational rate of 90% which corresponds to that of the proposed complex operational rate for normal stable operation.

4-3-1 Utility Requirements

The required capacities of the utilities are as shown in Fig. 4-1. The basic consideration and details of requirement are shown in 1-3, Part II of this report.

The main part of electric power is bought from outside power suppliers. Well water and sea water are utilized as industrial water and cooling water.

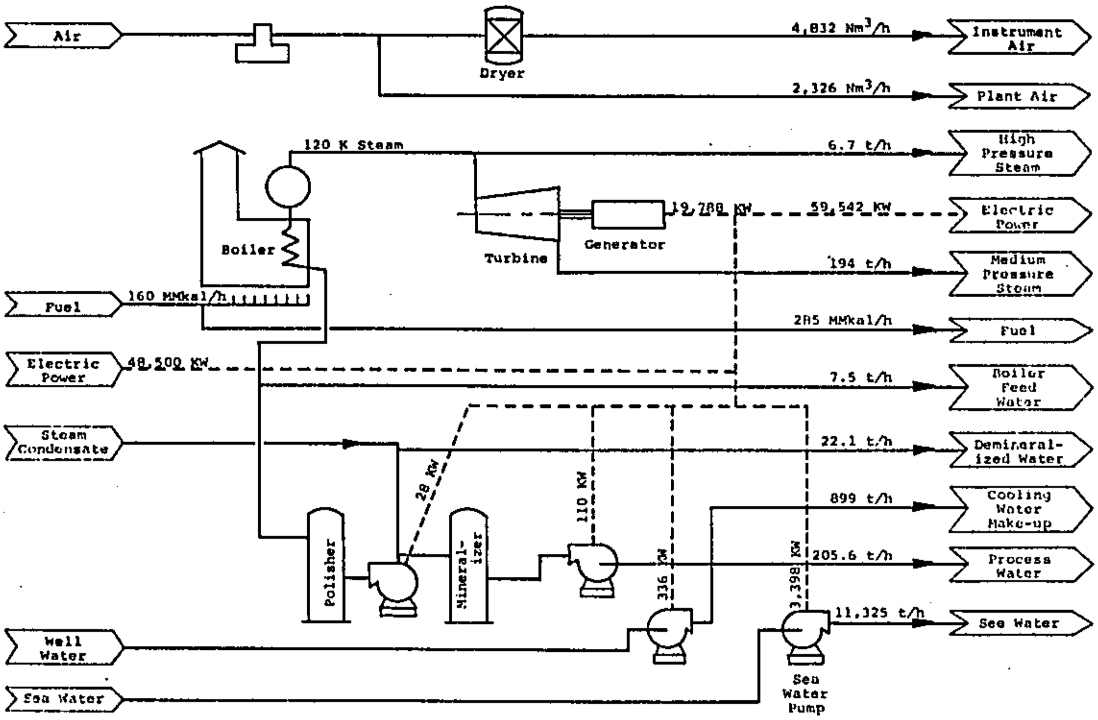


Fig. 4-1 Required Utilities and Capacity of Utility Facilities

4-3-2 Prices of Externally Supplied Utilities

(1) Electric power

The price of the power now supplied by MERALCO to Filoil Refinery is priced at US¢4/KWH on average. The generation is undertaken through thermo-power generation by firing fuel oil or crude oil.

Therefore, the price is rather high when compared with the prices of power distributed by NPC which is undertaking generation mainly at their hydro-electric power stations. Further, the Government is contemplating to convert the thermo-power generation to atomic power plants, by which the price of electric power will be lowered.

However, the position of atomic power plant is not clear at present, and the electric power supply to the Rosario area is now conducted by MERALCO. Therefore, the price of US¢4/KWH is taken as the basis and US¢6.0/KWH in 1980 is considered to be the externally supplied price with a 7% escalation per year.

(2) Well water and sea water

In some cases, the users of underground water must pay for the "right to using water"; however, in the Philippines no payment is required for the use of well water and sea water.

(3) Fuel

The price of fuel is set at US$10.95 in 1980.

4-3-3 Investment Cost Requirement

Construction cost estimations and requirements for the utility facilities are described in Chapter 3, Part III.

The construction cost requirement is summarized in Table 4-2. Total investment requirements are estimated in the manner as stipulated in Chapter 3, Part III.

Table 4-2 Basic Factors for Cost Estimation

	Construction Cost Plant Facilities	Construction Cost Off-site Facilities	Labor Requirements	Land Requirements
	x 10^3US$	x 10^3US$	Person	x 10^3m^2
Steam Generation	18,930	965	5	30
Electric Power Generation	24,039	3,746	4	1
Sea Water Intake	15,749	6,777	4	5
Well Water Facilities	454	858	2	2
Instrument Air Facilities	1,629	70	3	2
Plant Air Facilities	608	31	3	1
Oxygen and Nitrogen Generation	9,663	425	5	6
Water Treatment Facilities	4,517	189	3	3
	75,589	13,061	29	50

4-3-4 Prices of Utilities

The prices of utilities are estimated with a 10% ROI by using factors such as described in Table 3-6, Part I are resulted.

A summary of factors used in the calculation is as follows:

a. Depreciation: 15 years

b. Maintenance: 2% of ISBL cost + 1% of OSBL cost

c. Administration:

$$\left\{\left(\begin{array}{c}\text{Total Production Cost}\\ \text{- Contingency}\end{array}\right)+\left(\begin{array}{c}\text{Total Operating Cost}\\ \text{- Administration}\end{array}\right)\right\} \times 3\%$$

d. Labor requirement as stated in the organization chart (See 1-8, Part II)

4-4 Imported Products

For a comparative study of the price competitiveness of imported and domestic products and for a price projection of export products on the international market level, the prices of the items produced in the complex are estimated on the following assumptions:

4-4-1 Assumptions

(1) Basic prices

The basic prices of the Japanese products are determined by FOB Japanese Port, which will comprize the major portion of the total importation into the Philippines.

(2) Price escalation

Assumed prices of the FOB-Japanese-port-based products are estimated to increase by seven percent (7%) per annum up to 1980. This assumption is applied on the other items included in this study.

(3) Ocean freight

Ocean freight up to 1980 is assumed as follows:

1) Dry cargo

a.	LDPE	US$40/RT
b.	HDPE	
c.	PP	
d.	PVC	US$43/RT

2) Liquid

a.	VCM	US$34/RT
b.	Caustic soda	US$30/RT

(4) Marine insurance premium

a.	LDPE	2.5% on C&F x 110%
b.	HDPE	
c.	PP	
d.	PVC	
e.	VCM	0.8% on C&F X 110%
f.	Caustic soda	

(5) Dutiable value

Dutiable value at customs port is assumed to be CIF Manila plus point three percent (0.3%) of other charges.

(6) Terminal charge

1) Customs duty

Customs duty is calculated according to the current rates as given below:

a.	LDPE	30%
b.	HDPE	30%
c.	PP	30%
d.	PVC	50%
e.	VCM	10%
f.	Caustic soda	10%

2) Stevedoring charges

Stevedoring charges come in two (2) different options, either on a <u>flat rate</u> basis or on a <u>commodity rate</u> basis. The usual commodity rate is ₱6.39 per ton or "revenue ton" handled, while the normal prevailing flat rate is ₱7.00 per ton or "revenue ton" handled.

3) Arrastre

The prevailing charges for arrastre services is ₱9.00 per ton handled.

(7) Brokerage

The customs broker prepares all the necessary papers for the release of the cargo from the customs zone for a fee as stated below:

C&F/CIF Value of Shipments	Brokerage Fee
Shipment valued up to ₱10,000.00	₱100.00
Over ₱10,000.00 up to ₱20,000.00	₱150.00
Over ₱20,000.00 up to ₱30,000.00	₱200.00
Over ₱30,000.00 up to ₱40,000.00	₱250.00
Over ₱40,000.00 up to ₱50,000.00	₱275.00
Over ₱50,000.00 up to ₱60,000.00	₱300.00
Over ₱60,000.00	₱350.00

(8) Miscellaneous

The miscellaneous charges amount to ₱37.50. Since ₱300.00/shipment is needed to facilitate the release of the cargoes from the custom zone and the usual truck load is 8 tons/trip, it is assumed that one shipment is equal to one truck load.

(9) Customs charges

The customs charges incurred in the preparation of the release of the cargo are as follows:

1)	Forms/stamps	₱25.00
2)	Notarization	₱20.00
3)	Documentation	₱50.00

(10) Bank charges

1) L/C opening charges

Opening charge in L/C is point eighty-two percent (0.82%) of the CIF value of the goods.

2) TIR charges

TIR charge is four point six percent (4.6%) of the CIF value of the goods.

(11) Sales tax

The rate of the sales tax is seven percent (7%).

(12) Exchange rate

The predicted exchange rate by 1980 is ;

¥300 = US$1 - ₱8.12

(13) Delivery terms

Imported prices are based on ex-customs warehouse.

4-4-2 Imported Prices

The imported prices calculated according to the above assumptions are as follows:

(US$/MT)

	(A) FOB Japanese Port	(B) CIF Manila	(C) Ex-customs Warehouse	(D) (C)/(B)
LDPE	852	916.53	1,387.32	1.51
HDPE	966	1,033.67	1,560.23	1.51
PP	1,015	1,084.01	1,634.69	1.51
PVC	624	675.34	1,178.52	1.75
VCM	382	419.66	561.99	1.34
NaOH	127	158.38	233.07	1.47

4-4-3 Inference on the Price for Imported Products

Prices for Japanese petrochemical products would become the standards in the Southeast Asian countries as far as the period is taken for the future 5 - 10 years.

(1) Forecast on ex-factory prices of Japanese petrochemical products

The forecast on prices for a shorter period, say 1 - 2 years is usually done based on the supply demand balance of the products. Longer-term forecasts such as 5 - 10 years ahead are done by the production cost; and for the medium-term forecasts of 3 - 5 years ahead are done by combining the both.

For the purpose of forecasting the prices in 1980, a price forecast of a cost push type is made.

(a) Method for the price forecast

Ex-factory price for the petrochemical products can be divided into the following 4 portions:

a. Raw materials cost
b. Utilities cost
c. Depreciation cost for the facilities
d. Other expenditures

In regard to the raw materials cost, only the major materials were taken up. Auxiliary raw materials were included in the other expenditures, in which personnel cost, monetary interest, general administration and sales cost, and the profit are also included.

a) With respect to the raw materials cost, considerations were given to the price rising of crude oil occurred at the outset of 1974 and the same for naphtha from 1974 to 1975, and, thereafter let it made to slide on the commodity price indices.

b) Utilities costs were calculated based on the conversion into fuel, and fuel prices were let to slide to the crude oil price.

c) On the occasion of a newly made plant the construction cost has risen quite a lot, but during a period of about 5 years or so, Japan will still have majority of products produced by existing plants so that it is assumed that the depreciation cost will not change.

d) Other costs were presumed to slide onto the commodity prices.

(b) Naphtha price and commodity price rising rate in Japan

Naphtha price in Japan was made to ¥25,000/kl in October 1974, during which year the average price was ¥21,130/kl inclusive of the imported naphtha. However, the average price for the year 1975 is made to be ¥25,000/kl. After 1976 onward, it was assumed to rise up at a rate of 7%/y.

Rise up ratio of Japanese commodity prices was made to 60% for the period starting from 1971 down to 1975; and 7% from then on.

(c) Forecasting model for the ex-factory price of petrochemical products

a) Prices for ethylene, propylene and butadiene were calculated based on equivalent distribution formula.

b) Cost structure of each product was taken from that of 1971. Forecast models are as given below. Where, 'n' stands for 'zero' for the year 1975, 'r' stands for the commodity price rise up rate (7%).

(¥/Kg)

Ethylene	$P_E = 82,068(1 + r)^n + 8,247$
Crude Propylene	$P_{PC} = 49,241(1 + r)^n + 4,948$
Propylene	$P_P = 56,343(1 + r)^n + 10,175$
VCM	$P_{VM} = 73,919(1 + r)^n + 10,893$

LDPE	$P_{LE} = 152,371(1 + r)^n + 41,833$
HDPE	$P_{HE} = 183,284(1 + r)^n + 32,596$
PP	$P_{PP} = 180,541(1 + r)^n + 51,273$
PVC	$P_{PV} = 119,055(1 + r)^n + 20,207$

(d) Forecasts on ex-factory price

Price forecasts for 1980, calculated by the above models taken r = 7% are shown in Table 4-3.

Table 4-3 Ex-factory Prices of Petrochemical Products in Japan

(Unit: ¥/Kg)

	1975	1980
Ethylene	90.32	123.39
Propylene	66.52	89.20
VCM	84.81	114.60
LDPE	194.20	255.61
HDPE	215.88	289.74
PP	231.81	304.57
PVC	139.26	187.24

ANNEX

ANNEX I

POLLUTION CONTROL

ANNEX I POLLUTION CONTROL

CONTENTS

TABLES AND ILLUSTRATIONS

1. Pollution Control in the Petrochemical Complex

1-1 Waste Source in the Petrochemical Industry

The multiplicity of processes used and products produced by the petrochemical industry indicate the scope and complexity of the petrochemical waste problem. If one considers all of the process combinations used by the industry, it is obvious that many types of air and water pollutants can result. A general review of major pollutants in the petrochemical industry reveals that many of them can be traced to sources common to most processes.

(1) By-product formation

In some cases the by-products formed may have commercial value, but usually this is not the case and more often these products present waste disposal problems.

(2) Side-product formation

Side-products are compounds formed by reactions that are competing with primary reactions during petrochemical processing. These side-products are often isomers of the principal product. They may also be reaction products from impurities present in the feedstock or from other extraneous reactions which might occur during the petrochemical process. These side-products range in composition from pure chemical compounds which might have some commercial value to unrecognizable tars and complexes. Unlike by-product formation, they often can be controlled by modifying the conditions of the chemical reaction.

(3) Incomplete reactions

In petrochemical practice, no reaction ever goes to completion. The many reactions used by the industry vary markedly in their efficiency ranging from almost 100% completion to only 20% of completion. In other words, the product stream from any petrochemical process will contain varying quantities of the feed chemical which must be separated from the final product. Often, especially in cases of low conversion efficiency, the feed chemical is recyclable through the process; but in some cases, this is not economically feasible and they are disposed of as wastes.

(4) Mechanical and accidental losses

Every mechanical or physical operation is subject to various losses, including accidental leaks, spills, and explosions, or losses caused by human error. Mechanical losses include fluid losses from valves, defective seals in compressors and pumps, leaks in various units, etc. Volatile hydrocarbons are often lost to the atmosphere via relief valves and improperly sealed storage facilities. Storm water runoff from processing areas often contains pollutants which have seeped in from various process units.

The actual operations which are the principal contributors of water pollutants in the petrochemical industry are listed on the below.

One or more of these operations are used in each of the primary and secondary conversion processes and are usually the prime contributors of water-borne pollutants. All of these factors are involved in the waste discharge from a large petrochemical complex. The actual quantities and types of pollutants discharged by a plant will depend on several factors, including the type of raw petroleum feedstock used by the plant, the products manufactured, the quantity of in-plant control, the maintenance condition and types of processing equipment used.

(a) Steam distrillation
(b) Steam cracking
(c) Steam stripping
(d) Product washing with water, caustic, and acids
(e) Barometric condensing of steam-driven prime movers
(f) Catalyst regeneration
(g) Boiling and cooling water blowdown
(h) Rain water runoff from processing areas
(i) Cleaning operations (Process)
(j) Partial pressure reduction with dilution steam
(k) Transfer and storage operations
(l) Crude petroleum desalting
(m) Solvent processes
(n) Still tars from distillation processes
(o) Product and feedstock drying

1-2 Water Pollution Problems of the Petrochemical Industry

As mentioned above, petrochemical industry pollutants originate from the incomplete removal of the principal products or raw materials from the reactions, in the production of nonrecoverable or useless by-products, from equipment cleaning operations, and from such water uses as cooling and steam production. Waste water generation in the industry per unit of product varies so widely that an average value has little meaning except in a statistical sense. The principal contaminants in the industry's waste waters are BOD, COD, oil, suspended solids, acidit, heavy metals, color, taste and odor-producing compounds, and residual organic products and by-products.

The production of organic chemicals results in many types of contaminated waste waters; the treatment methods employed cover the range of known practical techniques. In-plant control is the first step in instituting treatment practices. Such controls include the salvage of unreacted chemicals, recovery of by-products, multiple reuse of water, reduction of leaks and spills, and changes in processing methods. These controls can result in reducing the concentrations of almost all potential pollutants. More importantly, they can reduce the volume of waste waters requiring treatment.

Physical treatment methods such as sedimentation of flotation are used primarily to remove coarse suspended matter and floating oils and scums. Filtration is used as a form of tertiary treatment for reuse and as a pretreatment for deep well injection. Chemical treatment is used primary as a pretreatment prior to sedimentation, filtration, or biological treatment. Biological treatment is most widely used in the industry due to the general susceptibility to biodegradation of the wastes, evidenced by relatively high BOD values.

Joint industrial-municipal treatment has proved to be very effective in treating organic chemical waste waters, particularly for smaller chemical plants located near large municipal treatment systems. Treatment costs play an important role in governing the expansion of joint participation. Rates established by municipalities vary widely. Where the municipal system is small and additional contributors would overload the treatment plant, the high rates are imposed to discourage industrial contributors.

The chemical industry has generally found that in-plant, separate treatment has economic advantages, particularly when significant quantities of contaminated waste water are involved. No significant percentage increase is expected in the amount of organic chemical waste waters that will be treated in joint systems in the near future.

The multitude of compounds present in petrochemical wastes results in a variety of adverse effects that may be exerted on the environment. A number of different methods have been used to classify these pollutional effects, and they are categorized as follows: (a) direct effects on the receiving waters; (b) effects on the usability of the water; and (c) physiological effects on the biological systems which either inhabit the receiving waters or use them for purposes which might render them subject to these effects. It should be recognized that ground water as well as surface water supplies can be subject to petrochemical pollution. Although the effects of pollution on surface water are more readily apparent, the less obvious pollution of ground water is equally important.

1-3 Phisiological Effect of Petrochemical Wastes

(1) Taste and odor

Petrochemical wastes contain many organic chemicals which can interact to produce tastes and odors. A study of a refinery waste water indicated that the neutral organic compounds consisting principally of aliphatic and aromatic hydrocarbons were the primary source of odor in the waste water.

The oxygenated and aromatic fractions of the neutral organic groups were also found to be the most odiferous in a similar investigation. These fractions were found to be a complex mixture of hydrocarbons, alcohols, aldehydes, ketones, esters, and nitriles; and the odors caused could not be attributed to any single compound. Many neutral organic compounds are commonly found in petrochemical wastes. Organic acids are also found, and many of these acids can cause tastes and odors in water.

Many taste and odor-producing compounds are therefore potentially present in petrochemical waste waters. Since tastes and odors in water have such an adverse effect on many uses, each waste must be carefully tested to determine if it causes such an effect and if so, what treatment is required.

(2) Toxicity

Toxicity includes the effects of pH, lack of dissolved oxygen, high temperature, and high dissolved solids as well as compounds which are classified as poisons. It is the poison effect which will be considered. The toxicity of a chemical is directly related to the dosage of the chemical ingested or otherwise taken into the body of the aquatic organism. In water, the toxic effects of a chemical can best be described in terms of the concentration threshold, below which physiological damage to the organism will not affect the subject during its normal life span, even if the compound is administered throughout the life span. This latter effect has been demonstrated for every toxic compound which has been studied, including the chemical carcinogens.

1-4 Reduction of Plant Waste Loads by Internal Improvements

An ideal method for controlling petrochemical pollutants is to minimize and control unavoidable losses near the source. Such practice reduces the cost of waste treatment and in many cases provides a valuable economic gain.

(1) Reduction of raw material losses

The losses of raw materials from storage, transport, and processing facilities are an important source of water pollution in the petrochemical industry.

For example, 0.21 percent of hydrocarbon feedstock may be lost in a carefully controlled plant, but in plants where less extensive control practices are employed, as much as 0.6 percent may be lost. Several improvements can be made by the industry to reduce the magnitude of these losses. The evaporation of light hydrocarbons from storage tanks can be controlled through the use of floating roof tanks and vapor recovery lines in tank vents and purge lines used for process start up and shut down can be reduced by installing refrigerated condensors ahead of the jets or by connecting the jet exhaust to vapor recovery systems.

Pipeline systems should be used to transfer raw materials whenever feasible in order to minimize transfer losses. Many petrochemical industries minimize transport hydrocarbon losses by either having subsidiary companies which supply raw materials or by producing their own. Probably the most important source of hydrocarbon raw material loss is from malfunctioning equipment lines, coupling leakages, pump gland leakage, etc. These losses can be corrected only by careful and persistent in-plant control.

(2) Recovery of usable reaction products

By-products represent a significant pollutional fraction of petrochemical waste waters and in many cases, by-product recovery from the process wastes is justified nor only in terms of producing a product, but also in reducing the pollutional load, the recovery of sulfur from crude oils, for example, minimizes the sulfur and mercaptan pollution. Many petroleum feedstocks contain large amounts of these sulfur compounds which must be removed to prevent the poisoning of process catalysts and products, and it is possible to recover sulfur from these hydrocarbon feedstocks by removing hydrogen sulfide in the form of acid gas using hydrodesulfurization. Hydrodesulfurization consists of removing the hydrogen sulfide from a process stream by stripping with steam or water. The hydrogen sulfide then is stripped from the water or condensate as acid gas.

Other sources of usable materials found in petrochemical wastes are the catalyst complex metals and the tars from catalytic processes. For example, a chemical company has sold a copper-containing tar to a smelter, which subsequently recovers the copper. Other metals such as nickel may also be present in evaluating quantities in tars and metal complexes. Chemical recovery companies also have reclaimed useful organic materials from tars. Usually the recovery of materials from these tars does not result in a direct profit to the petrochemical plant, but it may prove economically justified in terms of a reduced pollutional discharge.

Alkaline wastes from caustic washes are most significant and there are several methods available utilizing these caustic wastes which may prove more feasible than attempting to treat and discharge them. Some spent caustic solutions containing sulfides, phenolates, cresolates, and carbonates are marketable. Spent caustics containing large amounts of phenols and cresols can be sold to processors who separate and purify the cresylic acid fractions for commercial use. Spent caustics can also be regenerated by steam hydrolysis, electrolysis, air regeneration, and the use of slaked lime for reuse in caustic washing processes. While this method does not eliminate the problem of spent caustic disposal, it does minimize the quantity of spent caustic wastes produced. Steam hydrolysis is the most commonly used regeneration process and as much as 90 percent of the caustic soda can be restored by steam hydrolysis and air oxidation processes. Spent caustic and containing phenolic compounds can not be recovered using these processes, since phenolic compounds with sodium functional groups cannot be regenerated by steam hydrolysis. Phenolics also interfere with the oxidation process.

The recovery and recycle of process effluents containing unreacted raw materials allows substantial savings in raw material purchases and is common to most petrochemical processes in which the process reaction is incomplete. Many of the secondary reaction by-products are also valuable for use within the petrochemical plant or as marketable products.

Some of the possible uses for by-products produced in three common petrochemical processes are shown in Table AI-1-1. An example of the in-plant use of a by-product is the use of hydrogen chloride from hydrochlorination processes as one of the reactants in the additional chlorination of olefins. Hydrogen chloride can also be recovered and marketed as a salable product.

Table AI-1-1 Usable Side-products from Some Typical Petrochemical Processes

Primary Product	Side-Products	Use
Butadiene	Residue Gas (Hydrogen, methane, ethane, carbon dioxide)	Fuel
	Propane and Propylene	Feedstock for Ethylene, Alkylation
	Butane and Butenes	Recycle for Butadiene Manufacture; Feedstock for Alkylation
	Aromatics Oils	Resin or Plastic Manufacture
Ethylene	Residue Gas (Hydrogen, Methane)	Fuel
	Acetylene	Fuel for Welding Feedstock for Several Petrochemical Processes
	Ethane	Recycle for Ethylene Manufacture; Cracking Feedstock; Fuel
	Propane and Propylene	Propane Recycle for Eehylene Manufacture; Feedstocks for Several Petrochemical Processes (Alcohol, Alkylation, Polypropylene, etc.)
	Butane and Butylene	Feedstock for Synthetic Rubber Aviation Gas; Recycle to Cracking Process
	Aromatic Concentrate	Resin and Plastic Manufacture
	Heavy Oils and Tars	Refinery Charge Stock Fuel
Ammonia	Carbon Dioxide	Dry Ice, Bottled CO_2 Fuel Methanol Manufacture
	Helium	Lifting Gas Inert Gas
	Argon	Inert Gas

(3) Process modification

Process modifications incorporated into the design and operation of petrochemical processing units can often lead to reductions in the pollution output of the plant. These modifications can be classed as follows: (a) process selection, (b) prevention of product and chemical losses, and (c) modified operating conditions. The problems of waste control should be initially considered during the development and design of petrochemical processes, since they can often be an important factor in the economics of operation. Pilot plants used in the design of petrochemical processes can be sampled to estimate the process losses and the type of pollutants which will be discharged from the prototype unit. Analysis of these samples will indicate possible process modifications which can be implemented to reduce the pollutional load.

The incorporation of spent caustic regeneration schemes in process and plant design is an example of process selection used to reduce the waste load. The substitution of continuous processes for batch processes also tends to eliminate peak discharges of wastes and to reduce the cost of treatment required for the waste. Treatment facilities designed to absorb the effects of shock discharges of wastes through equalization basins or overdesigned tanks are considerably more expensive than those which are designed to treat a waste of relatively constant consumption.

The use of downgraded chemicals in processes which do not require high quality reactants can facilitate both process and waste control. This type of design utilizes the waste effluents from one process as reactants in another. In the ideal scheme, a chemical could be carried through several processes, each using a slightly lower quality reactant. Caustic washed are often so used in refining operations.

The hydrocarbon losses associated with the use of barometric condensers are an important source of pollution in petrochemical processes. The use of shell and tube heat exchangers rather than barometric condensers can significantly reduce this source of pollution. Process sampling taps connected directly to sewers are a source of pollution which can be avoided by returning drainage to the process unit rather than the sewer. Pollution caused by open bottom steam stills can be controlled by the installation of calandrias.

Distillation of waste streams can also be used to eliminate the loss of volatile materials. Furfural, for example, is recovered by azeotropic distillation, eliminating this compound from waste waters while producing a valuable solvent for use in extraction processes. One example of waste control by process modification has been arranging the interior of the pipe still to reduce friction losses through the trays and by lowering the temperature in the still, losses were reduced to one-six of their original quantity.

(4) Waste stream segregation

Segregation systems used by the petrochemical industry vary widely depending upon reuse requirements, effluent quality requirements, and possible undesirable effects which might occur if incompatible effluents are mixed. Three main segregated collection systems are normally used: (a) area drains which carry unpolluted cooling water and storm water runoff from uncontaminated areas; (b) a contaminated water system which contains process waters, polluted cooling waters, and storm water runoff from contaminated areas; and (c) a sanitary sewage system to collect plant domestic wastes. Many petrochemical plants segregate waste waters to an even greater extent. For example, oily process waste waters are frequently segregated from nonoily waste waters to facilitate oil removal through separators.

Segregation of many process streams may be necessary due to the incompatibility of certain waste components. Wastes containing cyanides, for example, must be separated from acidic effluents to prevent the release of gaseous hydrogen cyanide. Wastes with high solids concentrations are usually segregated from oily streams since suspended solids tend to increase the oil concentrations in oil separation effluents. Suspended solids can also interfere with the recovery of oils by increasing the solids contents of separator skimmings.

Most water reuse schemes also require additional waste stream segregation and the design of the segregation system will depend upon the type of water reuse being employed at the individual plant. If biological treatment is used to stabilize petrochemical wastes, the sanitary sewage is usually added to it prior to treatment as a source of biological nutrients; but many regulatory agencies require effluent chlorination when sewage is added, making such a combination unfeasible. The unpolluted cooling and storm waters which costitute the largest volume of water discharged by the petrochemical industry normally do not require treatment and are discharged directly to the receiving body of water or effluent waterway.

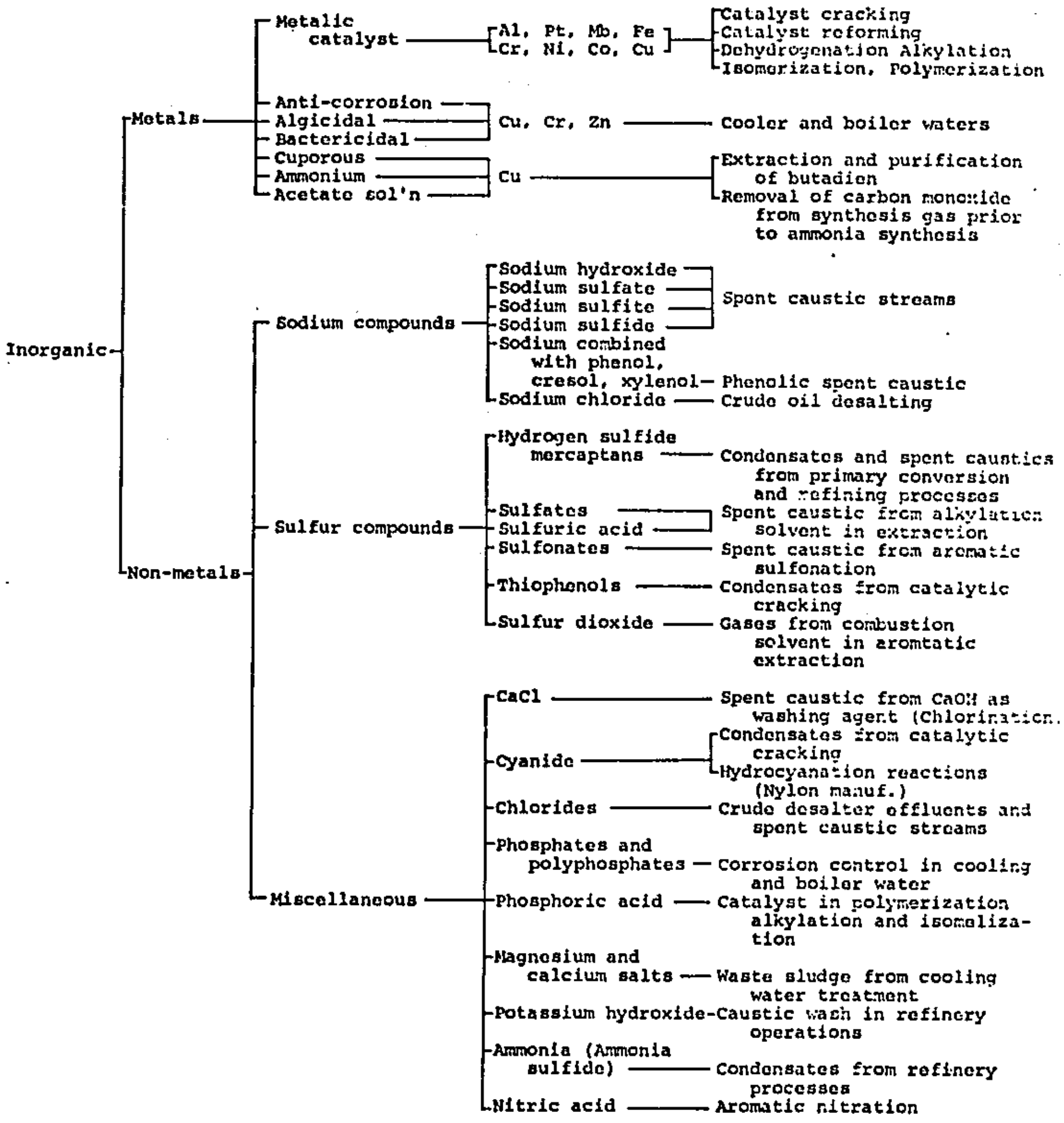


Fig. AI-1-1 Classification of Inorganic Compounds Which May Occur in Petrochemical Waste Streams

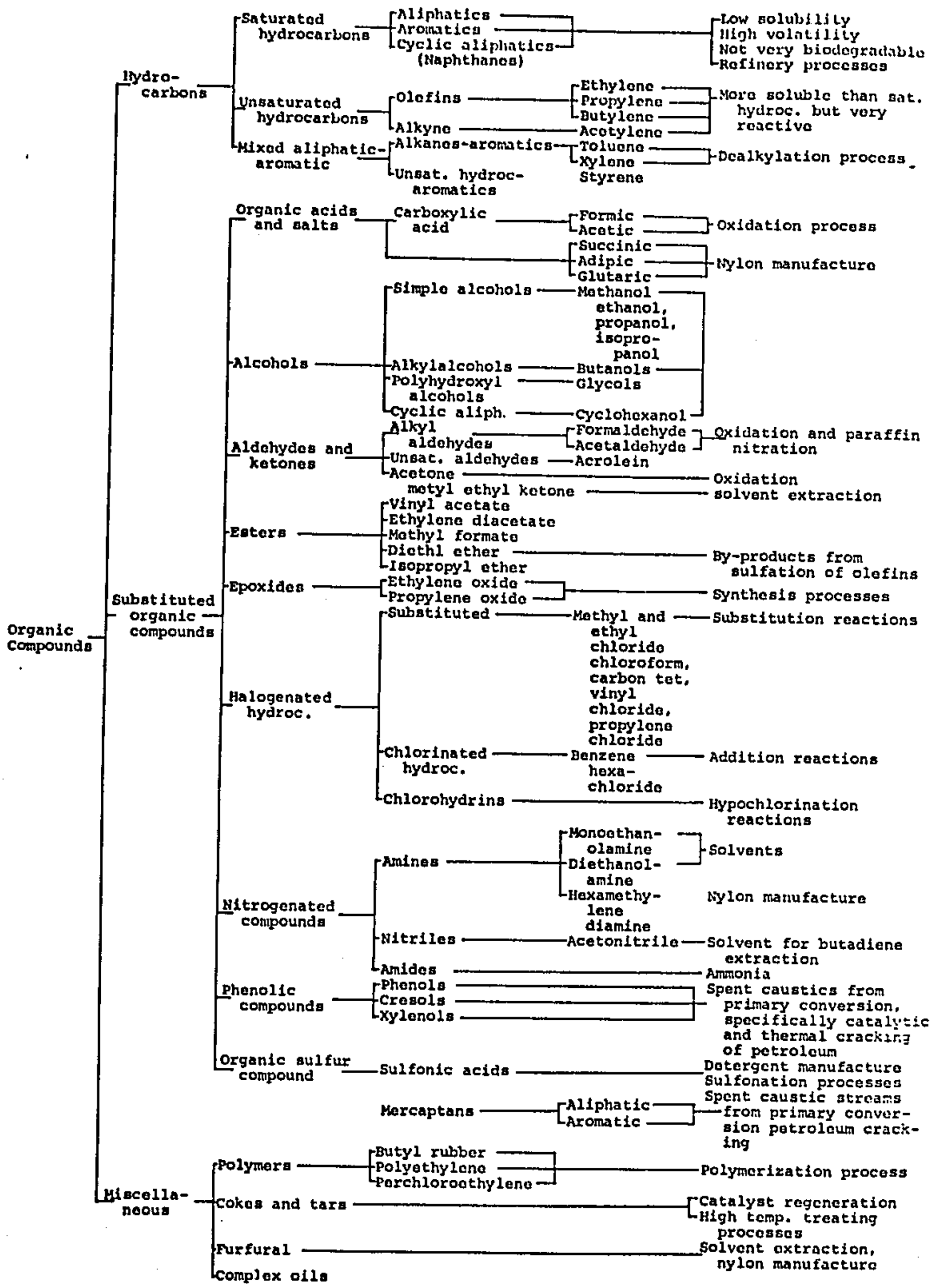


Fig. AI-1-2 Classification of Organic Compounds Which May Occur in Petrochemical Waste Streams

2. Guideline for the Method of Prevention of Mercury Poisoning

In the parliamentary report submitted by the Chemical Industrial Sub-committee of the Industrial Structure Examination Committee in February 1973, an emphasis was placed on the realization of harmonization between the chemical industry and the environment as the basic prerequisite condition for the industrial activities in view of the seriousness of the outcome of environmental pollution caused by chemical industry. As a means to promote such harmonization, development of the closed system is strongly stressed in the report.

Also by the Soda Sub-committee of the Committee, a report was presented to the Parliament emphasizing upon the elimination of pollution problems by means of improving the mercury cell electrolysis process by employing the closed system, by improving control technique, etc.

Based on the report, the Committee established a guideline in October, 1973. The outline of the guideline is as described in the following paragraphs:

2-1 Basic Policy for the Guideline

The basic policy for the guideline is that all the possible opportunities for the mercury leakage shall be prevented, and in the case of leakage of mercury from the electrolytic cell, an immediate recovery shall be effected. Contents of the method for the prevention of mercury poisoning are as follows:

(1) Waste water

Such a closed system shall be employed that contaminated water is strictly separated from uncontaminated water, and fresh feed water is minimized, and further the water which contacted mercury is not absolutely discharged.

a) Recycled utilization of contaminated water after total treatment

b) Reduction of the fresh feed water and application of an adequate treating method (for instance the evaporation method)

c) Separation of the contaminated water from the general discharge water such as cooling water, etc.

d) Treatment of rain water by means of effecting pavements, roof paintings, water reservoir vessels, or ponds

(2) Sludge (Salt-water mud)

The mercury discharge shall be reduced by means of cutting down the amount of the mud and mercury content in the mud.

The mud shall be treated as promptly as possible with the methods defined in the regulation.

a) Prevention of shifting of mercury into the mud

b) Reduction of the amount of the mud:

  - a. Utilization of high purity salt
  - b. Combination with the diaphragm cell process
  - c. Dissolution by acid

c) Elimination and recovery of mercury:

  - a. Centering
  - b. Extraction
  - c. Dissolution by acid

d) Disposal of waste mud after treatment

(3) Exhaust gases

The vaporization and dispersion of mercury shall be prevented by means of the following methods:

a) Tight sealing of the electrolytic cell and the auxiliary equipment

b) Tight sealing of the recycled salt water system

c) Elimination and recovery of vaporized mercury by aspirating from the electrolytic cells and the recycled salt-water system

d) Prevention of mercury leakage and prompt recovery of leaked mercury during the operation

e) Elimination and recovery of mercury contained in the discharged hydrogen gas

f) Decrease of the room temperature in the electrolysis room

(4) Products

The following methods shall be employed to reduce the mercury content in the products depending on the characteristics of the products being used as the basic raw materials.

a)	Caustic soda :	Elimination and recovery of mercury by filtration and adsorption
b)	Hydrogen :	Elimination and recovery of mercury by cooling, chemical treatment, adsorption, etc.
c)	Chlorine :	Prevention of mercury intrusion by using a mist separator

(5) Waste materials

Mercury contained in the waste material shall be thoroughly recovered, and at the same time, the treated material shall be reutilized as much as possible. When disposing of these materials, the treating method same as used for waste sludge shall be employed.

a) Consistent administration of the waste materials

b) Development of mercury recovering technology and improvement of recovering efficiency

c) Promotion of re-utilization and recycled utilization

d) Establishment of the technique of the waste treatment

(6) Others

The total amount of mercury within the system shall always be confirmed, and the mercury balance shall be well controlled.

a) Establishment of the operation manual and complete execution thereof

b) Reduction of the maintenance servicing frequency of the electrolysis bath and mercury handling frequency

c) Prevention of mercury leakage from the joints of the electrolysis bath and auxiliary equipment

d) Prevention of mercury leakage at the time of maintenance servicing of the electrolytic cell

e) Prompt recovery of leaked mercury (Paving at under-floor, construction of pits)

f) Establishment and accuracy improvements in the measuring technique of the total amount of mercury within the system

2-2 Practical Method

(1) Waste water

a) By means of maintaining the waste water channels and the water discharge systems, and by effecting a lining or coating to the channels, leakage of mercury through the channel walls shall be prevented.

b) If necessary, pits and water collecting devices shall be provided to be used only for the initial stage rain water treatment.

c) Provisions shall be so made as to recover and treat the rain water collected from the roofs of the electrolysis houses.

d) The water balance for the recycle-use portion shall be controlled by distinguishing the rain water, contaminated water, and general waste water.

e) Complete paving shall be effected to the places where the machinery and equipment are installed, and the areas in the vicinity of the plant in order to prevent the soil contamination.

f) The drain from the chlorine line and the hydrogen line (including the washing water for the direct-contact process) shall be recovered into the recycle system of the process.

g) Fluid dikes or fluid collection channels shall be installed around the solution baths, salt water bath, and other relative devices in order to treat the leaked salt water and the contaminated water.

h) The following provisions shall be taken regarding the recycled salt water:

   a. The balance in the circulating salt water shall be controlled by an adequate means such as the reduction in the feed water, the evaporation of the increased amount of the water, etc.

   b. PVC pipings shall be applied with adequate reinforcement, or the piping shall be replaced with reinforced plastics pipings.

   c. Leakage prevention shall be effected by mechanical seals of pumps and salt water pipings.

i) The sealing water at the top box, end box and the mercury pump shall be returned to the recycle system of the process, or shall be used as the thawing water.

j) Unless otherwise specifically required, the washing water, floor washing water, and other miscellaneous utility water used at the time of disassembly maintenance servicing of the electrolytic cell shall be returned to the process. The excess portion of such water shall be discharged after being treated.

k) As the pre-treatment for the contaminated water, pH adjustment, dechlorination by oxidation, filtration, etc. shall be effected to adjust the fluid properties. Thereafter, the contained mercury shall be treated by chemical processing, the activated carbon adsorption method, chelate resin adsorption process, ion exchange membrane cell process, and the combination thereof.

(2) Sludges (Salt-water mud)

a) The reduction and recovery of the contained mercury shall be conducted by means of the transfer-prevention method, or by the combination of the transfer-prevention method and extraction method.

b) If practicable, the reduction of the absolute amount of the mercury and the mud shall be intended by employing the HCl solution method or the sintering method.

c) The reduction of the generated mud shall be intended by the use of high purity salt, or by the combination of the high purity salt and the diaphragm cell process.

d) The minimization of contamination by the dispersion of mercury shall be effected by means of duly selecting the storage area of the untreated mud.

e) The final residue shall be stored and disposed in accordance with the relative regulations of industrial waste disposal.

f) The sludge other than salt-water mud shall be the subject of recovery of mercury as the waste material.

(3) Effluent gases

a) Tight sealing and recovery by aspiration shall be undertaken to the top box and the end box.

b) The mercury pump shall be of the canned type with a totally enclosed design or in the case of non-enclosed pump the aspirator employed for the gland part.

c) Sealing and aspiration shall be applied to the outlet of caustic soda of the thawing tower.

d) The aspiration units shall be well controlled.

e) The packing material for mercury piping shall be properly selected.

f) The temperature of the top box shall be reduced. (By means of cooling the mercury pumps and washing water)

g) The prevention of the vaporization and dispersion of mercury shall be prevented by means of reducing the mercury handling frequency and the electrolytic cell disassembly maintenance frequency.

h) When replacing the electrodes, the electrolytic cell shall be cooled before opening the sealing, and during the opening, a cover shall be placed over the cell, and the gas in the cell shall be aspirated.

i) The dispersion of mercury inside the electrolysis room by the effluent gas shall be prevented by the tight sealing of the electrolytic cell and the auxiliary equipment.

j) Especially when the salt-water process is in the state of alkaline salt water, tight sealing and aspiration facilities shall be considered.

k) The leakage of hydrogen gas shall be constantly checked, and when the leakage is detected the troubled part shall be properly repaired.

l) Regarding the hydrogen gas, both the purged and utilized portion shall be indirectly cooled, and then treated by absorption or adsorption of the contained mercury.

m) The air containing mercury shall be treated by feeding into the dechlorination tower in which the mercury is caught by the solution side (chlorine water) or adsorbed by the solution.

n) The aspirated gas shall be treated by the following methods:

   a. Hiponitrite soda-common salt method
   b. Potassium permanganate method
   c. Persulfuric ammonium method
   d. MRG method

o) A slight amount of mercury contained in the hydrogen gas shall be recovered by the following methods:

   a. Condensation separation method
   b. Adsorption method
   c. Absorption method

(4) Products

a) The caustic soda solution shall be treated by the activated carbon adsorption tower or by the activated carbon pre-coat filter.

b) The tank of un-filtrated caustic soda shall be closed type equipped with an aspirator.

c) When filtering the caustic soda, the points to be noted are the choice of the adequate filteration assisting agent, optimum filtration frequency, the improvement of the coating technique, adjustment of the filtration speed, etc.

d) Hydrogen shall be treated in accordance with the procedure for the effluent gas treatment.

e) The mercury contained in the chlorine shall be recovered by mist separation, cooling, and drying, and the washing solution shall be recovered into the salt water system.

(5) Waste materials

a) By means of reinforcing the material administration, the extent of waste material generation shall be minimized.

b) The generated waste materials shall be stored in a building with a paved floor, and shall be treated as quickly as possible.

c) When the waste materials are treated by dissolution, evaporation, distillation, sintering, etc., perfect effluent gas adsorption facilities shall be installed in order to prevent the air pollution and to maintain the wholesome working environment.

d) When solidifying the residue with concrete, the mercury concentration shall be reduced as much as possible.

e) The solid waste materials shall be treated by washing (with water, acid and arresting solution), mercury distillation, incineration, sintering and dissolution, and final residue shall be treated with the same procedure as for the mud.

f) The mercury contained in the mercury butter and electrolytic cell gas shall be recovered by distillation under reduced pressure and chemical treatment (acid treatment), and the final residue shall be treated by the same procedure as for the mud.

(6) Others

a) The mercury leakage from the joint parts of the electrolytic cell and the auxiliary equipment shall be prevented by welding, soldering and special packing.

b) The mercury leakage from the top box, end box and suction box of the pump shall be prevented by tight sealing of the covers or by the aspiration.

c) The mercury leakage from the drip receptacle of the caustic soda from the thawing tower shall be prevented by the method same as the above b).

d) The hold-up of the mercury shall be measured daily by direct measurement and shall be checked more than once a year by a radio-isotope measurement method in order to control the optimum mercury hold-up.

e) An ample slope in the shape of "V" shall be provided in the floor of the electrolysis room in order to assist the flow of mercury, and the surface of the floor shall be coated with resin or rubber without allowing any cracks.

f) The mercury on the floor shall be recovered by a vacuum cleaner, and water washing on the floor shall be minimized.

g) The mercury held in the pipings, holders and tanks shall be drained and recovered from time to time.

h) A mercury receptacle shall be provided at the necessary locations on the lower part of the electrolytic cell and the thawing tower, so that prompt mercury recovery can be effected at the time of leakage.

i) The location for the servicing of the electrolytic cell shall be well arranged for safe environment conditions.

j) At the time of disassembly maintenance of the electrolytic cell, the cell cover shall be opened after cooling and the cell shall be sealed with water.

k) The technique of measuring and detecting the mercury in the process shall be improved.

l) The possibility of changing the electrode from carbon to the metal shall be studied.

m) The following items related to administration, operation manual and education for operators shall be checked:

  a. The thoroughness of the promotion of the mercury cleaning intention on both management and labor

  b. The insufficiency in the workforce and facilities for mercury cleaning

  c. The enough preparation of the operation manuals for regular operations such as mercury butter cleaning, the semi regular operations such as electrode changing operation and irregular operations such as maintenance of the mercury pump

  d. The control of amount and quality of the water used for mercury recovery

  e. The record of the fresh feed and recovery of mercury.

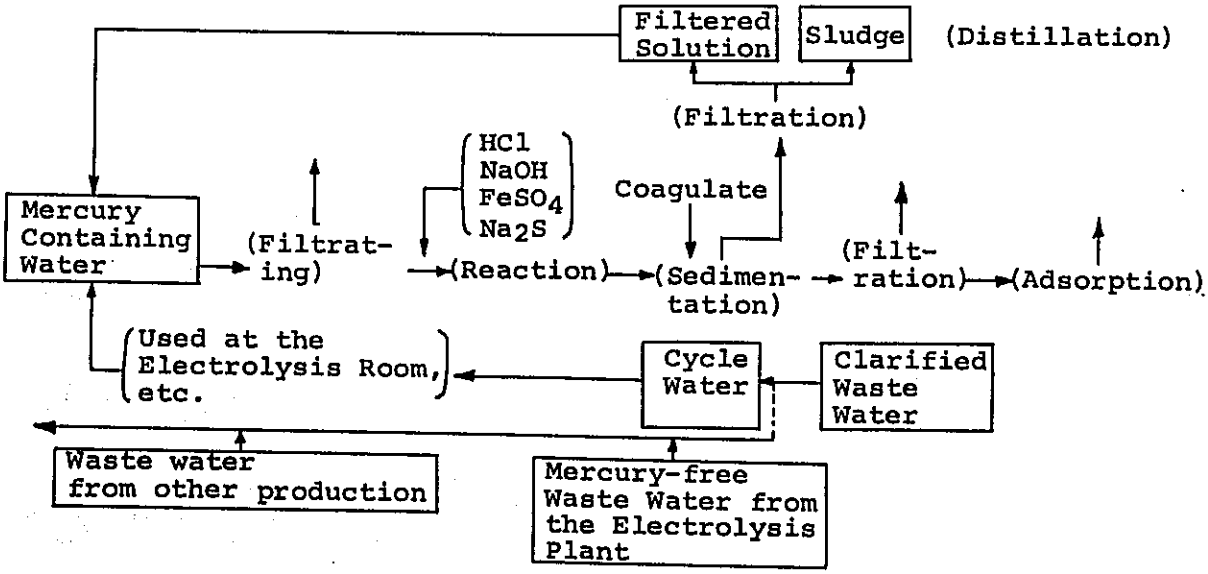
2-3 Reference

(1) Waste water

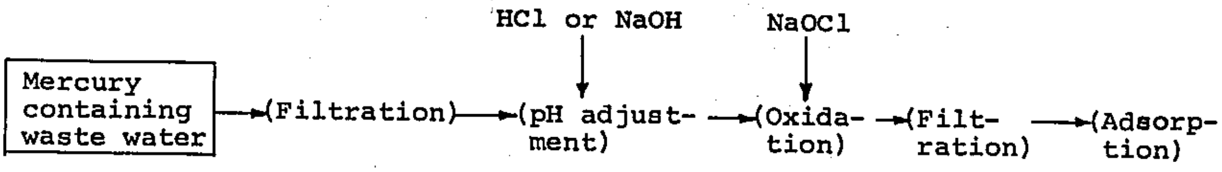
(a) The activated carbon adsorption process

Waste water is passed at a flow rate of 0.2 to 0.4 m/h through a filter pre-coated with powder activated carbon, or is flowed down at SV=3 to 4 into the filling tower filled with granular activated carbon.

(b) Sodium sulphate treatment, activated carbon adsorption process



(c) Chelate resin adsorption process



(2) Sludge (Salt-water mud)

(a) Shift prevention process, or combination of the shift prevention process and the extraction process

In the salt-water purification process, the mercury shift by adsorption to the precipitate shall be prevented by leaving an adequate amount of extricated chlorine (more than 10 mg) with control of a de-chlorine system or addition of Cl_2 and ClO^-.

Further, the mud slurry generated from the salt-water purification vessel is divided into solid and liquid by a filter press after the mercury is extracted by the hydrochlorite soda treatment.

(b) Hydrochloric acid dissolution process

Hydrochloride acid is added to the mud to control the pH of the mud at 1, then the acid soluble substances are dissolved by agitation. Thereafter, the pH is controlled to weak acid or neutral level, and the acid unsoluble substances are filtered.

The mercury contained in the filtrated solution is adsorped and treated with chelate resins and ion-exchange resins. By means of this treatment the total amount of generated mud and the mercury contained in the mud will be reduced.

(c) Roasting process

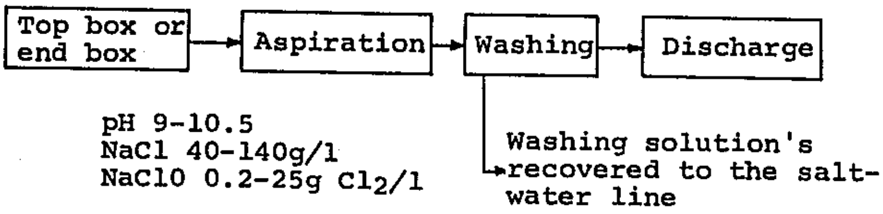
The mud is heated to 700 to 1,200°C by means of indirect heating by a heavy oil burner or electric heater in a rotary kiln or on a stationary bed, and the generated mercury vapor is recovered by cooling. By this treatment the mercury content in the roasted residue will be reduced to less than 3 ppm.

The rotary kiln has an advantage of preventing the emergence of blocks of the mud because of its uniform roasting and adequate heat conductivity.

(3) Exhaust gases

Following figures of a) - d)

(a) Sodium hypochlorite - brine process



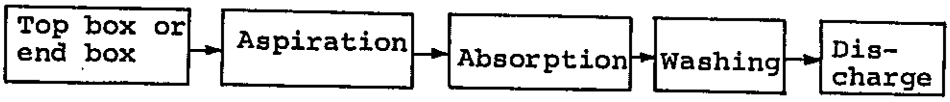
pH 9-10.5
NaCl 40-140g/l
NaClO 0.2-25g Cl_2/l

Washing solution's recovered to the salt-water line

(b) Potassium permanganate process

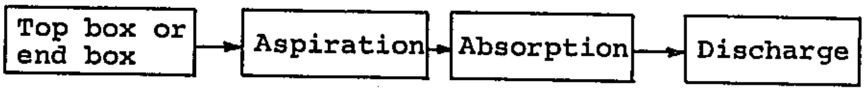
$KMnO_4$-H_2SO_4 Solution

(c) Ammonium persulfate process



$(NH_4)_2S_2O_8$-H_2SO_4 Solution

(d) MRG process



(4) Products

(a) Caustic soda

a) Pre-coat filter (example)

i) Activated carbon

o Type : Charcoal o Activation : Steam process

o Grain distribution				
	-60 mesh	10%	150-200 mesh	18%
	60-100 "	23"	200-250 "	11"
	100-150 "	9"	250- "	29"

ii) Treating conditions

o Pre-coat amount 300-500g/m^2 o Filtration speed Approx 0.26 m/h

b) Adsorption tower (example)

i) Activated carbon

o Type : Charcoal o Activation : Steam process

o Grain distribution

12-20 mesh	4%	48-60 mesh	33%
20-35 "	29"	60-80 "	10"
35-48 "	23"	80- "	1"

ii) Treating conditions

o SV = 1.0 m/h o LV = 1.0 m/h

(b) Hydrogen

a) Basic process (elimination of mercury from hydrogen)

Hydrogen from the electrolysis → (Cooling) → (Elimination) → for use of vent

b) Cooling (first stage elimination)

i) The mercury and drain generated from the thawing tower is condenced and separated by indirect cooling (using cooling water or a coolant), and most of the mercury is recovered.

ii) The separated mercury and drain are recycled to the thawing tower for reuse.

c) Elimination (elimination of a slight amount of mercury)

i) Condensation separation method:

The partial pressure of mercury in hydrogen is reduced by compression or subzero cooling or their combination, and the mercury is condensed and separated.

Crude gas → (Compression) → (Water cooling) → (Brine cooling) → Refined gas

ii) Adsorption method

The mercury vapor is adsorbed and separated by an adsorbent such as activated carbon or a molecular sieve.

iii) Absorption method

The mercury vapor is absorbed and separated by contacting the hydrogen gas with adequate washing liquids in a packed column or a spray column.

ANNEX II

RECENT ASPECTS OF BASIC TECHNOLOGY IN PETROCHEMICAL INDUSTRY

ANNEX II RECENT ASPECTS OF BASIC TECHNOLOGY IN PETROCHEMICAL INDUSTRY

CONTENTS

TABLES AND ILLUSTRATIONS

1. Recent Trends of the Electrolysis Processing of Salt

1-1 Introduction

The environmental pollution caused by waste has spurred the development of a process for electrolysis of salt to replace the mercury cell process and the converting of present mercury cell operations. In the United States of America, the diaphragm cell process has generally been adopted because of the advantage of direct feeding of brine pumped from salt mines and because of the proximity to Canada where excellent asbesto is produced. On the country, in the Europe and Japan, the mercury cell process is popular because it produces superior quality products such as caustic soda, chlorine, etc., at the present.

However, the employment of the mercury cell process will be leagally prohibited for new plant installations in all countries. Furthermore, even in countries such as Japan, laws have been passed so that even the operation of existing plants employing the mercury cell process will be prohibited in the near future.

The abolishment of the mercury cell process has brought with it severe problems regarding the quality of products, especially of caustic soda, which has a wide scope of applications.

Since caustic soda from the diaphragm cell is low in concentration and includes impurities such as salt, it requires evaporation and rectification depending on its intended applications.

In this paper, we will first deal with an outline of the technical aspects of each type of electrolysis process and will include the development present situations, as follows:

a) Mercury cell process and countermeasures for mercury leakage

b) Diaphragm cell process and purification of caustic soda

c) Ion exchange menbrane cell process and its situation of development

Secondly, the process related production and application related consumption in the U.S.A. and Japan will be surveyed. The break down of the consumption is reported for high purity caustic soda (equivalent to the mercury cell products). The required amount of high purity caustic soda for the Philippines is estimated on the basis of the above data and by the demand for caustic soda projected by the BOI of the Philippines.

The improvement of the existing mercury cell plants will be a vital problem in the future for pollution control in the Philippines. As a reference for the above problem, "Guideline for the Method of Prevention of Mercury Poisoning" is reported in Chapter 2, Annex I.

1-2 A Comparison of the Electrolysis Processes and Their Situation of Process Development

1-2-1 General

Although the future employment of the mercury cell processes is generally regarded as almost impossible, intensive research and development on the closed system have significantly decreased mercury leakage. In comparison with the 200 to 300 gr/caustic soda ton of the actual recorded mercury loss before, a newly constructed plant can reduce mercury loss to the level of 1 to 2 gr/caustic soda ton.

The diaphragm cell process does not use mercury, so the pollution problem due to the mercury does not exist. However, evaporation and the purification of caustic soda, depending on the application, are necessary. Consequently, a considerable amount of capital investment for the facilities and the consumption of steam (av. 4 t/t) is required. Therefore, only large-scaled production units can economically employ the diaphragm process. High purity caustic soda is produced mainly through the ammonium extracting process.

The ion exchange membrane cell process is now under commercial development and can in principle, produce high purity caustic soda without using the mercury. Accordingly, the need for evaporation and purification facilities is expected to be eliminated or reduced in the future. Because of this, private companies and governmental organizations connected with this field paying great attention and expending strong efforts on the developing of the compact plants with low operating costs which employ the ion exchange membrane cell process. At the present, however, the only actual commercial plant in operation was developed by Asahi Chemical and has been operated since April in 1975. It was reported that the Hooker plant would start operation of the 275 st/d plant for production of 10% concentrated caustic soda in April, 1975.

The quality of caustic soda produced commercially by the ion exchange membrane process has been proved to be suitable for the most stringent industrial requirements. For example, the textile manufacutures association of Japan has assured its suitability for the rayon production.

However, the actual record of operation regarding the life of the ion exchange membranes, mechanical troubles, and plant economics has not been established at present.

1-2-2 Mercury Cell Process and Countermeasures for Mercury Leakage

The mercury cell process has the following advantageous characteristics:

a) High purity caustic soda

b) The evaporation and purification cost are not necessary for caustic production

c) Comparatively low construction cost

Togetherwith the above advantages, the following unfavorable characteristcs are present.

d) Mercury is consumed for electrolysis, and the leakage of mercury causes severe environmental pollution

e) Electricity consumption is comparatively high.

Since the environmental pollution problem due to the mercury leakage is the most serious, and potentially fatal, issue, the employment of the mercury cell process in future plant construction is regarded as impossible for any country. Therefore, countermeasures for mercury leakage in existing plants are the most important considerations in the improvement of plant process, operation, and design and engineering. Regarding practical methods for prevention of mercury pollution, "Guideline for the Method of Prevention of Mercury Poisoning" is shown in Chapter 2, Annex I. In this paper, the prevention of mercury leakage by employing closed system is described. A typical process flow diaghram of the mercury cell process is shown in Fig. AII-1-1.

Actual records of the mercury loss from electrolysis plants are shown in Table AII-1-1, which gives the break down of the mercury loss through leakage channels. The record shown in (1) is a typical example of the mercury loss which was observed in plants before the mercury pollution was not regarded as a serious problem. The record shown in (2) gives the effects of intensive imporvement in mercury leakage prevention which was conducted by Hoechst - Uhde, in Germany, in a newly constructed electrolysis plant. The record shown in (3) is the latest performance of the electrolysis plant of Kashima-Denki, Japan, reported in May, 1975.

Table AII-1-1 Actual Record of Mercury Leakage in Electrolysis Plant

	Actual Record (1)		Actual Record (2)				Actual Record (3)	
			Before the Processing of Hg Leakage Prevention		After the Processing of Hg Leakage Prevention			
	gr/NaOH-ton	%	gr/NaOH-ton	%	gr/NaOH-ton	%	gr/NaOH-ton	%
Chlorine Gas	-	-	-	-	-	-		
Hydrogen Gas	6	2	9.68	62.1	0.0026	0.076	2	100
Caustic Soda	10	4	2.64	17.0	0.399	11.6		
Waste Brine Mud	134	54	2.2	14.1	2.2	64.2		
Washing Water Waste Water	56	28	0.35	2.3	0.123	3.6		
Ventilating Air of Electrolytic Room Miscellaneous	31	17	0.7	4.5	0.7	20.5		
Total	237	100	15.58	100	3.43	100	2	100

Table AII-1-2 outlines the allowable leakage of mercury in several countries. The comparison of Table AII-1-1 with Table AII-1-2 shows that the possible amount of mercury leakage from newly constructed plants under the control of present regulations, will be considerably lower than the mercury loss which caused the severe environmental problems of the past.

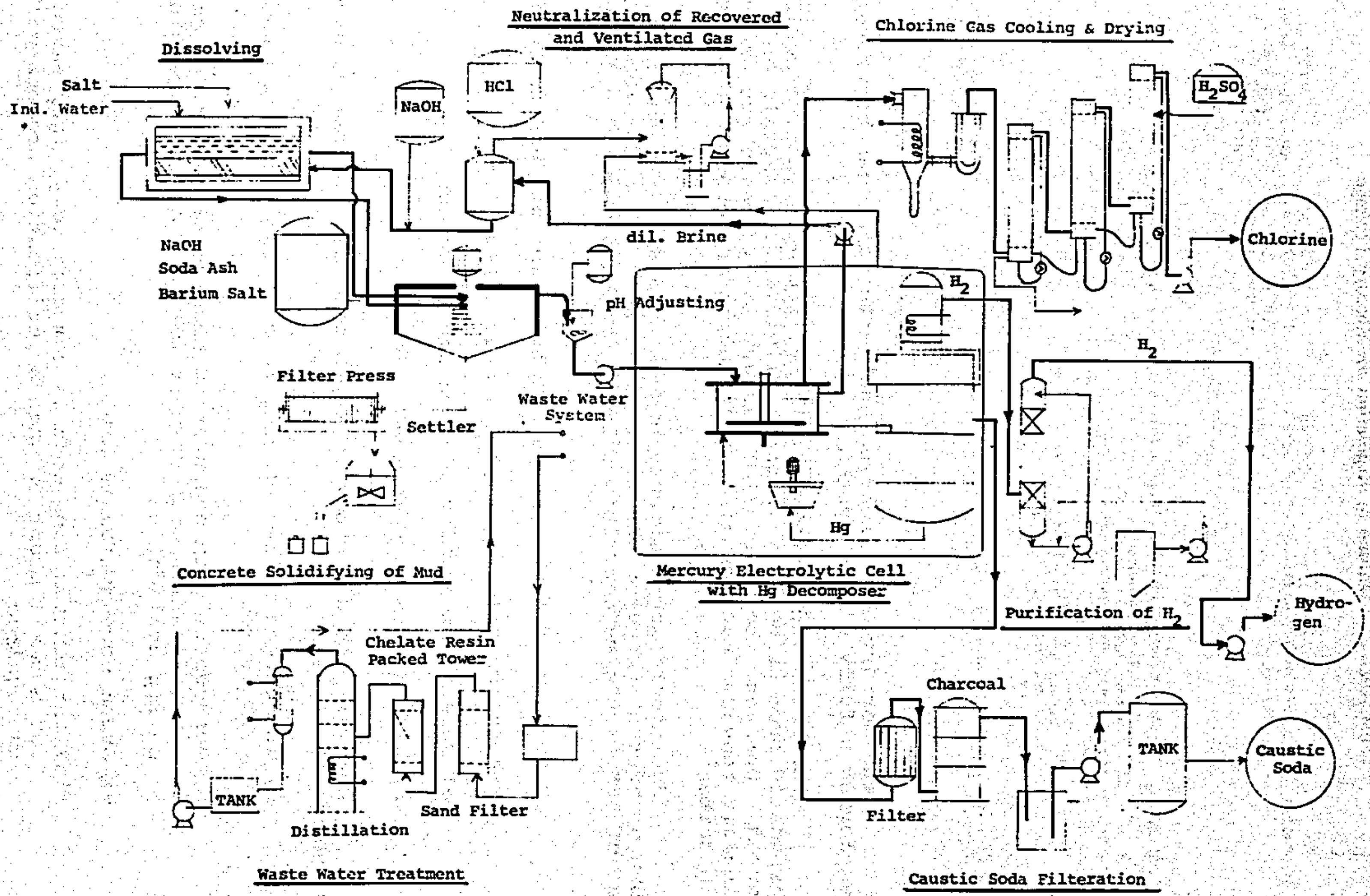


Fig. AII-1-1 Process Flow Scheme of Closed System Mercury Cell Process

Table AII-1-2 Outline of Regulation on Mercury Leakage

	U.S.A.	Canada	Sweden	Japan
Atomospher	2300 gr/day-plant 0.1 mg/m^3	0.05 mg/m^3	1.0 gr/Cl_2-ton 0.5 gr/Cl_2-ton in Hydrogen	After 1978, mercury cell process is inhibited for operation
Waste Water	450 gr/day-plant For existing plants after 1977 Max. 0.28 gr/Cl_2-ton Av. 0.14 " For newly constructed plants after 1977 Max. 0.14 gr/Cl_2-ton Av. 0.07 "	2.25 gr/Cl_2-ton	0.5 gr/Cl_2-ton	
Solid Waste Disposal	No regulation			Reserve by solidifying the residual mud

The mercury loss in brine and waste water can be prevented by employing a fully closed system.

The brine and waste water is recirculated and the residual is separated from brine. It is then solidified by concrete and reserved. However, the removal of mercury in hydrogene gas and ventilating air requires much caution in operation and design of electrolysis units.

By integrating a mercury process and a diaphragm or anion exchange membrane process, the mercury leakage can be lowered to the level of 65% of the mercury loss from an ordinary mercury cell process.

1-2-3 Diaphragm Cell Process and Purification of Caustic Soda

Diaphragm cell process has the advantages of:

a) No environmental pollution due to mercury

b) Electricity consumption is low

However, the diaphragm cell process has the following negative characteristics:

c) The concentration of caustic soda generated from the cell is low

d) Consequently an evaporation process is necessary, and a purification process unit may be required depending on its application

e) Steam consumption is high

f) Plant construction cost is comparatively high, especially for a small plant or when a purification process unit is required.

A typical diaphragm cell process flow diagram is shown in Fig. AII-1-2, using the process integrating ammonium extractive purification process. In the diaphragm cell process, the caustic soda solution generated in the cell is 12 to 15% concentration and contains 15 to 20% salt.

Table AII-1-3 shows the comparison of qualities of caustic soda between the diaphragm process and the mercury process. The solution is condensated through vaporizing water by a multi-effective evaporator, utilizing steam, and the salt is separated by crystallization through cooling. The concentrated caustic soda contains generally 1 to 1.2% of salt, 0.1 to 0.3% of sodium chlorate ($NaClO_3$), iron, etc. The main purposes of purification are the removal of salt, iron, and decoloration. The ammonium extraction process is generally used in the U.S.A. and the crystallization process and the complex salt process are applied. The first two processes are especially effective for the removal not only the salt also the other impurities.

Table AII-1-3 Comparison of Quality of Caustic Soda (Liquid)

	Diaphragm Process	Mercury Process
NaOH	48 - 50 %	49 - 51 %
NaCl	1.0 - 1.2 "	0.002 - 0.005 "
Na_2SO_4	0.02 - 0.1 "	0.001 - 0.003 "
$NaClO_3$	0.1 - 0.3 "	0.00008 - 0.0001"
Na_2CO_3	0.1 - 0.3 "	0.04 - 0.05 "
Fe_2O_3	10 - 50 ppm	Fe 1.5 - 3 ppm
SiO_2	50 -200 "	2 - 5 "
Ni	0.1 - 0.5 "	-
Cu	-	0.2 - 0.3 "
Hue	Light Color	-

(1) Ammonium extraction process (D-H process)

Impurities such as salt, sodium chlorate, etc., are eliminated through extraction by the liquid ammonium absorbent, and the efficiency of separation is remarkable (Ref. Fig. AII-1-2). Furthermore, the operability is satisfactory due to liquid phase handling. The facilities for the process unit consist of high pressure equipment employing higher grade metals such as nickel, so the process is suitable for large-scale production. Table AII-1-4 shows the example of analyses of caustic soda before and after the purification treatment.

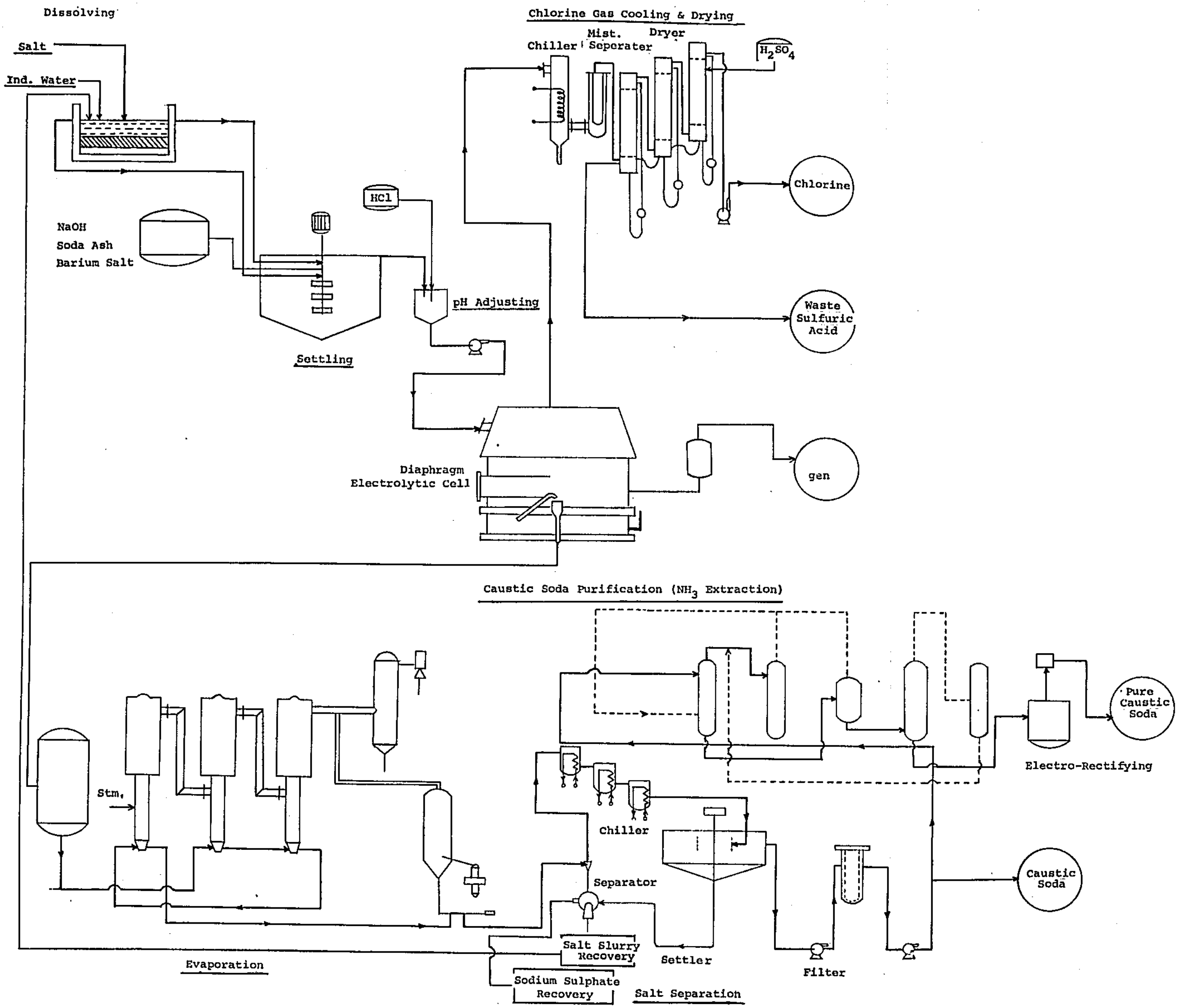


Fig. AII-1-2 Process Flow Scheme of Diaphragm Cell Process with Caustic Purification Unit

Table AII-1-4 Example of Analysis

	Before Purification	After Purification
NaOH (wt.%)	50.5	50.5
$NaClO_3$ (%.S.B.)	0.3	0.0004
NaCl (". ")	2.0	0.05
Fe (ppm. ")	7	2
Ni (" . ")	2	0.2
Pb (" . ")	4	0.4
Cu (" . ")	0.2	0.1
Turbidity	Muddy	Clear

(2) Crystallizations process (Hydrate process)

After the crystallization of caustic hydrate by cooling the low purity caustic soda solution, the high purity caustic soda solution is produced through melting the caustic hydrate separated from mother liquor. Although the disadvantage of this process is the repeating of cooling and heating, separation of almost all of the impurities is possible. The actual operations were recorded by I.G. and Hooker from 1930 through 1955, however, this process was once abolished at the time of conversion to either the mercury process or the ammonium extraction process. Recently, Tokyo Industrial Testing Center and Aginomoto Co., have independently developed a new crystallization processes by pursuing vast improvements in the old process. The process developed by TITC employs a direct refrigeration system using a freon for the crystallization.

(3) Complex salt process (Sodium sulphate process)

By adding sodium sulphate to the caustic soda solution, the salt is removed through precipitation as an insoluble complex salt, i.e., $NaOH \cdot NaCl \cdot Na_2SO_4$ or $4NaOH \cdot 4NaCl \cdot 5Na_2SO_4$. As the purified caustic soda contains 0.2 to 0.3% of salt and also contains sodium sulphate, this process is not employed at present.

1-2-4 Ion Exchange Membrane Cell Process and Its Situation of Development

Ion exchange membrane process is commercially under development for the production of high purity caustic soda without using mercury. At present, a few commercial plants such as Asahi Chemical are using this process. On the basis of reported data and published information, the established ion exchange membrane process has the following characteristics:

a) No environmental problem due to mercury

b) The purities of caustic soda and chlorine generated in the cell are high

c) Consequently, the caustic soda purification process unit is not necessary

d) In comparison with the diaphragm cell process, the overall plant cost can be lowered by the reduction in the capacity of or need for evaporators, and boilers, etc.

e) The contribution to the saving of energy is large.

Because of these characteristics, many private companies and governmental organizations are showing great interest or using intensive efforts to develop this process.

(1) Outline of process

The ion exchange membrane cell process employs a consistent cation-ion exchange membrane instead of a porous diaphragm and separates sodium ions, through selective electrodialysis, to produce high purity caustic soda with little salt content. In Table AII-1-5 , examples of the analysis of caustic soda are shown and list the products from the ion exchange membrane process, the Asahi Chemical process and other processes. For the analysis of the gaseous products, refer to Table AII-1-6.

Table AII-1-5 Example of Caustic Soda Analysis

Process	Ion Exchange Membrance	Mercury Cell	Diaphragm Cell
NaOH	24.8	48.7	50.5
Na_2CO_3	0.035	0.04	0.07
NaCl	0.0022	0.003	1.0
Na_2SO_4	0.0013	0.0013	0.02
$NaClO_3$	0.0015	0.0001	0.04
SiO_2	0.0034	0.004	0.02
FeO_3	0.00015	0.00015	0.0009
Al_2O_3	0.0008	0.001	0.0005
CaO	0.0004	0.0003	0.003
MgO	0.0001	-	0.0006
Mn	0.00002	-	-

Table AII-1-6 Analyses of Gaseous Products in Various Processes

Process Component		Ion Exchange Membrance*	Mercury	Diaphragm
Caustic Soda	NaOH	17 - 25	49 ± 1	49 ± 1***
	NaCl	0.01**	0.005	1.0 - 1.5***
	$NaClO_3$	-	-	0.2 - 0.5***
Chlorine	Cl_2	99.5	98.5	96.5 - 98
	H_2	0.03	0.2 - 0.5	0.1 - 0.5
	O_2	0.2	0.2	1.0 - 2.5
	CO_2	0.3	0.4	0.1 - 0.3
Hydrogen	H_2	99.9	99.9	99.9

* Actual record of the pilot plant of Asahi Chemical

** Concentration at 49% NaOH

*** Concentrated and desalted value

To establish an effective commercial operation, the following technical targets should be attained:

a) Costs of investment, operation, and maintenance for the ion exchange membrane process should be equivalent to those for the other processes

b) The caustic soda solution from the cell should contain at least 20% of caustic soda, and should not include salt and chloride

c) Life of ion exchange membrane	8,000 hr or longer
d) Current efficiency	90% or greater
e) Electricity voltage	45 volt or less
f) Current density	30 A/dm^2

To satisfy the stringent conditions listed above, the further development has been conducted on the ion exchange membrane, an anode, and electrolytic cell.

1) Ion exchange membrane

The durability of the membrane is one of the most critical issues for commerical operation. The membrane must withstand the difference of pressures between the anode section and the cathode section, hot temperatures, hydraulic loads at the inlet of the electrolyte, and the corrosive atomospher of the chlorine, hypochlorous acid, etc. It must also endure against the abscission of polymer chains by oxidation, the desorption of the ion exchange radicals, and the decreasing of strength of the membrane due to its swelling.

Furthermore specialized design and construction are required for the supporting frame of the membrane, as well as for the development of the membrane itself. Nafion, (perfluorosulfonic acid type cation-ion exchange membrane) which was developed by Du Pont, is regarded as a prospective material at present. The structual formula is as follows.

$CF_2\text{-}CF_2\text{-}CF_2\text{-}CF\text{-}(\text{-}O\text{-}OF\text{-}CF_2\text{-}CF_2)\text{-}O\text{-}CF_2 \cdot CF_2SO_3Na$

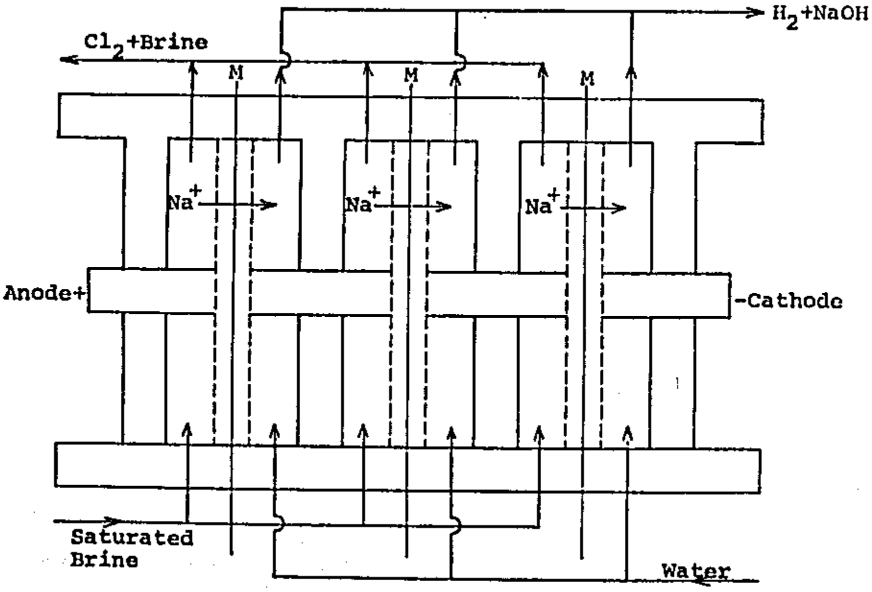
Nafion has a high secondary transition point and is durable in a hot and oxidizably corrosive environmental atomospher. Several years of testing show a prospect of durability of more than 5 years. However, the material requires more improvement to establish 90% or greater current efficiency under hot temperatures.

2) Electrode

Metalic electrodes have the characteristics of low electrode potential, little heat generation, high current density, and they prolong the life of a diaphragm or a membrane. Regarding anodes, an expanded titan electrode, coated with platina and rhodium, has been developed. De Nora has developed the expanded steel electrode which is coated by a rare metalic oxide as a cathode.

3) Multi-electrode type electrolytic cell

A compact multi-electrode type electrolytic cell has been developed which decreases the voltage drop of the circuits, thereby enabling all the electric circuits to work at the same voltage. Schematic diagram is shown in Fig. AII-1-3.



Note: M; Membrane

Fig. AII-1-3 Schematic Diagram of Multi-electrode Electrolytic Cell

(2) Development situations of the ion exchange membrane process

Since the development of the metalic electrode in 1970, and of the fluoric high durability ion exchange membrane in 1971, the development of this process has progressed rapidly. In this paper, the development situations are outlined for the companies in corresponding fields in the U.S.A. and Japan (Ref. Table AII-1-7). The U.S.A. originated the basic technology for this process, such as the fluoric membrane. In Japan, the mercury process will be completely prohibited in 1978, so intensive efforts are now being made among many companies to establish commercial plants using the ion exchange membrane process.

Table AII-1-7 Development Situation of Ion Exchange Membrane Process

Company	Situation
Diamond Shamrock (U.S.A.)	A pilot plant of employing Nafion membranes is under construction, and will start operation at the beginning of 1976 to gather commercial operation data. The development is aimed at producing 35-45% NaOH in the large sized cell which will be more economic than the existing diaphragm and mercury process by eliminating evaporation process unit.
Hooker (U.S.A.)	It was reported that a commercial plant was to begin operation in April, 1975 in Tacoma. The plant is planned to increase the capacity to 275 st/d maximum. Also, a plant having a capacity of 45 st/d is under construction and will start in November or December in 1975. The concentration of caustic soda for both plants is 100 gr/l, but Hooker regards 20% caustic soda as possible.
PPG (U.S.A.)	Currently testing in a small-scaled pilot
Olion (U.S.A.)	Currently developing 5-8 st/d cells
Ionic (U.S.A.)	Currently selling FOROMAT systems (small-scaled electrolysis units for production of bleaching agents, chlorine, and caustic soda for water treatment, etc.)
Asahi Chemical (Japan)	A commercial plant of the capacity of 40,000 mt/y, employing Nafion membrane has been operating since April in 1975. The quality of NaOH, 25% concentration, was tested in July, 1975 by the Association of Chemical Fiber Manufactures of Japan and was found to be suitable for rayon production.
Asahi Glass (Japan)	A pilot plant employing a membrano which was developed by Asahi Glass, is under construction and will begin operation on September, 1975. It is reported that the process produces high purity caustic soda having the content of 40% or greater, doer not require an evaporation unit, and the electricity consumption, 2,200-2,500 KW per ton of caustic soda, is lower than that of the mercury process, 2,500-2,600 KW, and the content of salt is 0.03 percent which is equivalent to that of the mercury process.
Maruzene Petrochemical	Currently developing an ion exchange membrane and ion exchange membrane process
Tokuyama Soda	A pilot plant is under construction and will begin operation on October, 1975, to establish the technical background, until the end of 1977, for commercial plants. The process will be developed employing both Nafion and an own-developed membrane.
Toyo Soda	A pilot plant will begin operation at the end of 1975. The other situation is the same as that of Tokuyama Soda.

Except for a few cases such as Asahi Chemicals, the ion exchange membrane cell process is not employed in commercial plants. Even for the Asahi plant, which started in April, 1975, the period of actual operation is very short. It seems likely that at least one year of operation will probably be necessary to confirm the operability, reliability, economy, etc. of the commercial plants.

In the U.S.A., the most advanced group (Diamond Shamrock, Hooker, Ionics) has appeared to require two years to prepare for the commercial supplying of the electrolytic cell in order to produce high purity and high concentration caustic soda with large capacity and high efficiency. At present, it is generally accepted that the ion exchange membrane process will be advantageous for small-scale plants producing comparatively low concentrated caustic soda when high purity caustic soda is required.

The progress of process development of individual companies may be divided into three categories:

a) Construction of commercial plants producing middle concentrates (25%) high purity caustic soda, spurred by governmental regulation abolishing the mercury cell process

b) Construction of pilot plants for the further development of the processes for producing highly concentrated caustic soda

c) Construction of commercial plants producing low concentrated (10%) caustic soda for internal company consumption or for special uses.

1-3 Specifications for Caustic Soda

For the reference of the specification according to intended use of caustic soda when the conversion of the electrolysis process is required, the industrial standard specification in Japan and examples of specification in the U.S.A. are shown as follows:

For chemical fiber production:

Examples in the U.S.A.	Table AII-1-8
JIS K 1204 (Japan)	Table AII-1-9

In the U.S.A. the industrial code for the practice of the analysis is established such as ASTM, but it remains under the mutual agreement between the suppliers and the consumers.

For industrial use:

JIS K 1203 (Japan)	Table AII-1-9

For edible use:

Codex Food Additive (Japan)	Table AII-1-9

Table AII-1-8 Quality Specification of Caustic Soda for Rayon in the U.S.A.

Rayon Maker / Component		Company A	Company B		Company C
			P - D*	M	
NaOH	(%)	50 - 52	50 - 52	50 - 52	50 - 52
NaCl	"	0.015	0.05	0.004	0.03
Na_2CO_3	"	0.10	0.15	0.05	0.40
Na_2SO_4	"	0.004	0.01	0.003	0.07
$NaClO_3$	ppm		4	1	
SiO_2	"	20	150	15	200
Al_2O_3	"		10	5	30
Fe	"	2	6	3	5
Ni	"		0.3	0.3	1
Cu	"	0.1	0.3	0.3	
Mn	"	none	0.5	0.3	0.4
CaO	"	6		5	30
Cr	"			0.5	

Note: P-D: Purified Diaphragm Process Chlorine
M: Mercury Process Chlorine
*: Specification of mixed use

Table AII-1-9 Quality Specification of Caustic Soda in Japan

Codex Food Additive (for edible application, liquid)

NaOH	95-120% of Indication
SO_4	0.2% > (Na_2SO_4 0.3%>)
Na_2CO_3	2% >
AS_2O_3	2ppm >
Heavy Metal	(pb) 30ppm >
Hg	0.05ppm >

JIS K1203 (liquid) (as of 45% NaOH)

	1st Grade	2nd Grade	3rd Grade	4th Grade
Na_2CO_3	1 %>	1 %>	1 %>	1 %>
NaCl	0.1 "	0.5 "	1.3 "	1.6 "
Fe_2O_3	50ppm	100ppm	200ppm	300ppm

JIS K1204 (for chemical fiber, liquid) (as of 45% NaOH)

	1st Grade	2nd Grade
Na_2CO_3	0.3 %>	0.4 %>
NaCl	0.07 "	0.40 "
SiO_2	0.03 "	0.15 "
Al_2O_3	0.02 "	0.04 "
CaO	50ppm>	80ppm>
Fe_2O_3	30 "	50 "

1-4 Demand for and Production of High Purity Caustic Soda

The future requirements for high purity caustic soda, equivalent in quality to the product of the mercury cell process in the Philippines, are estimated on the basis of the intended use forecast by the Philippines BOI, and an analysis of the actual consumption records in the U.S.A. and Japan.

The analysis mentioned above has been developed from reports provided by the Japanese Government and the related industries noted in introduction. In May, 1974, the Philippines BOI forecasted the intended use demands during 1974 through 1977 in the Philippines.

1-4-1 The U.S.A.

The breakdown of the caustic soda demands is shown in Table AII-1-10. The 1974 minimum requirements for high purity products and diaphragm process products are shown in Tables AII-1-11 and AII-1-12 respectively. As shown in the following table, the diaphragm cell process is popular in the U.S.A., therefore, the use of the diaphragm process product is also popular, covering 90% of the total consumption of caustic soda. Countermeasures for the problems involved with using the diaphragm process product are generally taken, for example, the installment of stainless steel equipment to guard against $NaClO_3$ corrosion. The percentages of production by individual process, of caustic soda are as follows:

	Caustic production percentages
Diaphragm process	71.8%
Mercury process	24.6%
Melting and non-electrolysis process	3.6%
Total	100.0%
Total production capacity	5,000,000 mt/y

Table AII-1-10 Breakdown of Demand of Caustic Soda in the U.S.A.

Application	(%)	Fraction of High Purity Caustic Soda (%)
Chemicals	40 - 45	12 - 14
Paper. Pulp	12 - 14	-
Alumina	5 - 8	-
Textile Processing	3 - 6	-
Rayon. Staple Fiber and Cellophane	4 - 5	100
Soap. Detergent	3 - 4	-
Petroleum	3 - 5	-
Export	8 - 10	-
Others	10 - 15	-
Total	100	10

Table AII-1-11 Application of High Purity Caustic Soda in the U.S.A.

Application	(10^3st/y)	(%)
Rayon and Cellophane	500	4.5
Chemicals. Intermediate	270	2.5
Farmaceuticals. Reagents	220	2.0
Ion Exchange Resin Regeneration (concentration of uranium etc.)	110	1.0
Total	1,100	10.0

Table AII-1-12 Application of Diaphragm Process Caustic Soda in the U.S.A.

Application	(10^3st/y)	(%)
Photosensitive Materials	100	0.9
Food Products	200 - 300	1.8 - 2.7
Others (Pharmaceuticals, reagents, etc.)		
Total	300 - 400	2.7 - 3.6

Furthermore, the purification facilities having the capacities of 15 to 20% of total capacity are presently equipped, however, the operational rates of these units are low due to the existence of the supply of high purity caustic soda from mercury process. The cost for purification is reported 10 to 20 US$/t in the U.S.A., but there is no record of using only the purified diaphragm process caustic soda for rayon production.

1-4-2 Japan

The breakdown of the caustic soda demands and minimum requirements for the high purity products in 1973 are shown in Tables AII-1-13 and AII-1-14, respectively. The possibility of using diaphragm process products are assessed on the basis of the results of inquires* to members of the associations of manufactures who use caustic soda in Japan.

Note: * Technical Sub-committee of Japanese Caustic Soda Industrial Association, <u>Survey report on the use of caustic soda</u>, May, 1975.

Table AII-1-13 Demand Actual Record of Caustic Soda in Japan

(10^3ton)

	1972 Demand	H. P. Caustic Soda Demand	Fraction of H. P. Caustic Soda (%)
Chemical Fiber	282	282	100
Paper Pulp	186	24	(13)
Cellophane	51	35	68
Alumina	213	6	(3)
Dyeing	59	2	(3)
Chemical Industry			
Detergent	57	-	-
Dye Intermediate	139	22	16
Inorganic Chemicals	651	127	20
Soda for Electrolysis	88	-	-
Organic. Petrochemical	247	81	34
Others	378	47	13
Total	2,351		
Flavoring Material*	88	52	60
Petroleum Refinery	24	-	-
Others	479	46	11
Total	2,944	724	24.6

* MSG 83%, Amino acid soy bean souse 14%

Souce: Survey by Caustic Soda Manufactures Association

Table AII-1-14 Consumption Actual Record of Caustic Soda for Inorganic Chemicals in Japan

(10^3ton)

	1972 Demand	H. P. Caustic Soda Demand	Fraction of H. P. Caustic Soda (%)
Sodium Silicate	105	-	-
Sodium Sulfite	156	17	10.8
Sodium Tripoyphosphate	97	30	30.9
Bromine	16	-	-
Sodium Dichromate	263	80	30
Glauber's Salt			
Sodium Cyanide			
Sodium Hypochlorite			
Others			
Total	636	127	20

Source: Survey by Caustic Soda Manufactures Association

The requirement for high purity caustic soda is estimated at 23% of the total caustic soda demand. Regarding the production of seasoning, especially mono-sodium gultaminate (MSG), the following problems are predicted for using the diaphragm process caustic soda; the deterioration of the products due to the increase of salt and Fe compounds, etc., the detrimental effects on the fermentation process and the corrosive effects on the equipment due to the existence of salt and hypochlorite, and the increase of loss in rectifying process and the deterioration of waste liquid treatment capacity due to the existence and the accumulation of salt.
The ratio of the production capacities according to the processes in 1974 is as follows:

Diaphragm	4.8%
Mercury process	95.2%
Production capacity total	3,200,000 mt/y

1-4-3 The Philippines

The caustic soda company production capacities in 1975 are shown in Table 3-9, Chapter 3-4, Part II, which is surveyed by the BOI of the Philippines. At present almost all caustic soda is produced through the mercury process, but all the future plans of electrolysis plants employ diaphragm process. The total production capacity of caustic soda in the Philippines in 1974 is 61,800 mt/y.

The projection of intended use demands of caustic soda during 1974 through 1977 made by BOI is shown in Table 3-7, Chapter 3-4, Part II. On the basis of above, the minimum requirement of high purity caustic soda in 1977 is estimated by taking the correlations derived from the actual consumption records of the U.S.A. and Japan. As shown in Table AII-1-15, the estimated total requirement of high purity caustic soda in 1977 is 14,800 mt/y. The caustic soda demand for textile industry is limited to the materials for the mercerizing process, etc., not for the production of rayon, so it seems unnecessary to use high purity caustic soda. For the production of MSG, the high purity caustic soda seems necessary due to the previously stated reason.

Table AII-1-15 Demand Estimation of High Purity Caustic Soda in the Philippines

(t/y)

Industry	Caustic Soda* Demand 1977	Fraction of High Purity Caustic Soda (%)	High Purity Caustic Soda Demand
Soap & Detergent	30,484	-	-
Pulp & Paper	24,506	13	3,190
Textile	14,039	-	-
Mining	10,500	-	-
Oil	985	-	-
Silicate	729	-	-
Mono Sodium Glutaminate	16,123	83	11,650
Power Plants	1,631	-	-
Petroleum	460	-	-
Total	99,457		14,840

* BOI Estimation

2. Problems in Gas Oil Cracking

2-1 Introduction

In the countries where no domestic hydrocarbon resource is available, the petrochemical feedstock is limited to petroleum product which are delivered from refineries. In this case, utilization of heavy fractions is sometimes requested because the naphtha fraction can be used as crude gasoline which is most profitable.

The above situation can apply to the case in the Philippines. No hydrocarbon resource is found at present and gasoline is the main product of refineries.

Presently the heavier feedstock is attracting interests and efforts in the development of the pyrolysis technology among the related plant owners and engineering companies.

The raw materials for the olefinic petrochemical complex, must be selected by taking into account the technical reliability of the processing as well as the effective utilization of hydrocarbons, and the economics of plant according to the feedstock employed.

For the practical purposes, the actual profitability of the plant can not be established without the sufficient background in the technical aspects.

In this regards, this paper describes briefly the problems faced with the utilization of heavier feedstocks, and outlines the presently applied technologies. In addition, discussion is made regarding limitation in the pyrolysis of heavier fractions.

At present, the use of vaccum gas oil as a feed for an ethylene plant is likely to be unfavorable in view of both plant economy and technical reliability in vaccum gas cracking.

2-2 Difference in Properties between Naphtha and Gas Oil

The most important property of a feedstock for potential ethylene production is its hydrogen content. The higher the hydrogen content, the greater the yield of ethylene. Generally speaking, a fraction having a lower boiling point (lower molecular weight hydrocarbon) has a higher hydrogen content than a heavier stock, and consequently gives a higher yield of ethylene (Ref. Table AII-2-1).

The second most essential property is the type of compounds involved. Both naphtha and gas oil are complex mixtures containing many hydrocarbons: paraffins, isoparaffins, naphthenes, and aromatics. The normal paraffins are the most favorable for the production of oleffins. Isoparaffins and naphthenes produce lower ethylene yields. Practically all the aromatics remain unconverted in the pyrolysis reaction, naphtha, generally difined as a material boiling up to 200°C, is produced from topping units. An atmospheric gas oil can be broadly defined as a material having boiling point above 200°C and is produced from topping unit, i.e., an atomospheric distillation unit. Vaccum gas oil, which has experienced severe thermal hysteresis during the course of production, inherently contains many aromatics and few paraffins (Ref. Table AII-2-1). Therefore, for all the practical purposes, the vaccum gas oil yields considerably lower ethylene (Ref. Table AII-2-1).

Table AII-2-1 Examples of One through Yields from Fractions of Kuwait Crude

	(c) Full Range Naphtha	(b) Atm. Gas Oil	(c) Vacuum Gas Oil
Sp.Gr, 60/60°F	0.713	0.833	0.904
Boiling Range °c	32 - 171	292 - 327	370 - 470
Hydrogene, wt%	15.2	13.7	12.4
Aromatics, wt%	7	24	46
Severity	High	High	High
Yields wt%			
Methane	15	13.7	9.1
Ethylene	31.3	26.0	16.6
Ethane	3.4	3.0	4.4
Propylene	12.1	9.0	13.3
Butadienes	4.2	4.2	4.1
C_4, S	2.8	2.0	5.8
C_5 to 190 - 200°c	22.0	20.6	17.3
Cracked fuel oil	6.0	19.0	27.3
H_2,C_2H_2, C_3H_4, C_3H_8	3.2	2.5	2.1*
	100%	100%	100%
Data Source;	S & W**	S & W **	Exxon ***

Note: (*) includes C_4 acethylenes
(**) The data seemed pilot furnace records.
(***) Commercial plant data.

2-3 Sulfur in Feedstock

Unless an adequate pre-treatment is conducted, generally the content of sulfur in the feedstock increases as the feed becomes heavier. The allowable content of sulfur in the feed for an ethylene plant is approximately 0.2 wt% because of mechanical troubles and deterioration of the by-products which occur when higher concentration are present. Therefore, the cost of desulfurizing should be accounted for in the price evaluation of the feedstock.

2-3-1 Sulfur Content in By-products

As shown in Table AII-2-2, in the cracking of the gas oil, increased the sulfur in the feed is mainly concentrated in the by-product liquids, especially heavy cracked fuel oil. Therefore, 60% to 70% of the sulfur in the feed is eliminated, it is possible to burn the cracked gas fuel oil internally without causing sulfur dioxide emission.

Table AII-2-2 Effect of Sulfur in the Feedstock

Kg per 100 Kg of ehtylene	Naphtha	Atm. gas oil
Sulfur in feed	0.175 (0.05%)*	0.652 (0.15%)*
Sulfur in:		
Hydrogen Sulfide	0.14	0.15
Pyrolysis gasoline	0.017	0.03
Cracked fuel oil (200°c + 340°c)	0.018	0.06
Cracked fuel oil (340°c plus)		0.412

Source : Lummus Co.

Note: (*)% Sulfur content in feed

2-3-2 Sulfur Corrosion

Although most of the sulfur compounds in the heavier feedstock remain unchanged and a small part of them crack to hydrogen sulfide, its concentration of is quite high. Hot sulfurization is most evident in the range of 600°C in the cracking coil and transfer line exchangers. The concentrated hydrogen sulfide in the condensated water produces serious partial acidic corrosion on the equipment surface during the course of the cracked gas compression.

2-4 Plant Cost Difference

Because the ethylene yield becomes lower as the feedstock becomes heavier, an ethylene plant employing heavier feedstock will handle a larger amount of materials in the process.. Therefore, the plant cost becomes greater as heavier feedstocks are used. The approximate plant cost rations according to the feedstock employed are shown in Table AII-2-3.

Table AII-2-3 Relative Investment Cost (Inside Battery Limit)

Feedstock	Investment Cost Ratio
Ethane	1.0
Propane	1.14
Naphtha	1.24
Light Atmospheric Gas Oil	1.34 - 1.43
Heavy Gas Oil	1.39 - 1.49

Source: Lummus Co.

2-5 Yield Pattern

As described in the previous section, the yields of olefins and aromatics become lower as the feedstock becomes heavier, the pyrolysis of the heavier feedstock requires a larger amount of feedstock and yields a larger amount of by-products in liquid phase, especially fuel oils (Ref. Table AII-2-1).

The prices of the feedstock and of the by-products greatly affects the viability of a plant which pyrolyzes the heavier feedstocks. Therefore, it is difficult to determine the most profitable raw materials. These depend on the individual configulation of price structure of by-products such as fuels, relating to the specific project.

However, in the Philippines, since specific tax is exempted, the prices of the petroleum fractions such as naphtha, gas oil, etc., are not greatly different. Furthermore, the possibility of utilizing the by-products is slight, therefore they must be evaluated on a fuel basis. This effects the economy of an ethylene plant adversely when the heavier feedstocks are used.

Another problem is the increase of the cracked fuel oil boiling over 300°C which is a by-product for approximately 50% to 100% of the ethylene production, in contrast with 6% to 20% in the naphtha cracking. In the gas oil cracking, cracking furnace must be designed to burn the cracked fuel oil due to the shortage of the cracked fuel gas. The cracked fuel oil boiling over 300°C has a much higher viscosity than that by-produced in the naphtha cracking. Therefore, more difficulties are expected in the burning of the craked fuel oil uniformly required for an effctive and stable pyrolysis operation, as well as in the transportation and storage.

2-6 Technical Difficulties in Gas Oil Cracking

In addition to the problems accompanying with gas cracking discribed previously, there are several technical difficulties in the selection design, and construction of equipment for a favorable commercial operation. The formation of cokes and the generation of tars significantly increases, and the yield of ethylene significantly deteriorates, as the feedstock becomes heavier. Therefore, more progress in engineering technology will be required in order to conduct stable and continuous commercial operation of thermal cracking furnaces and quenching systems, e.g., transfer line heat exchanger.

At present the pyrolysis of the heavier feedstock is not common, but trial commercialization, especially in the pyrolysis of the vaccum gas oil, is beginning in several companies. Therefore, the engineering technology employed for the heavier feedstock pyrolysis is being developed on the basis of a more popular technology in naphtha pyrolysis. In particuler, concerning the heater tube type thermal cracking, the establishment of favorable operation is being attempted through combining existing equipment and selecting appropriate conditions in the design and operation, rather than developing to fully new equipment.

Since these are the present circumstances, this paper outlines the present situation of technology in pyrolysis, including naphtha cracking, and discusses the possibilities of applying them in the heavier feedstock pyrolysis.

2-6-1 Thermal Cracking Furnaces

Since the epochal development of short resisdence time type furnaces, SRT, by Lummus, the high selectivety thermal cracker has been rapidly developed by employing vertical tubes and high inflow of heat to the pyrolysis coils. Stone and Webster have since developed an ultra reverity cracker, USC, employing comparatively small pyrolysis coils to obtain a high yield of ethylene. Furthermore, KTI (formerly Selas) recently began to build compact furnaces employing a stringent marginal design based on engineering advances and successful experience in the horizontal crackers.

The three companies mentioned above have 70 % or more of the thermal crackers for ethylene plants in the world. Presently, the pyrolysis of heavier fractions such as gas oil is attracting enthusiastic attention in related fields worldwide. In the development of heavier feedstock cracking furnaces, the technology developed for naphtha cracking is mainly employed. The following Table AII-2-4 gives the outline characteristics of the thermal crackers developed by the companies mentioned above.

In addition to the information in Table AII-2-4, Mitsubishi Petrochemical Industry and Mitsubishi Heavy Industry has developed a new type cracker employing oval coils. The oval coils increase thermal inflows through increasing the surface area of the cracking coils, without sacrificing the economy of furnace, by the specific arrangement of coils and burners. Furthermore, Exxon has succeeded in the commercial pyrolysis of vaccum gas oil in Port Jerome, France, through on stream water decoking of coils in 1 - 2 week intervals.

Table AII-2-4 Characteristics of Representative Thermal Crackers

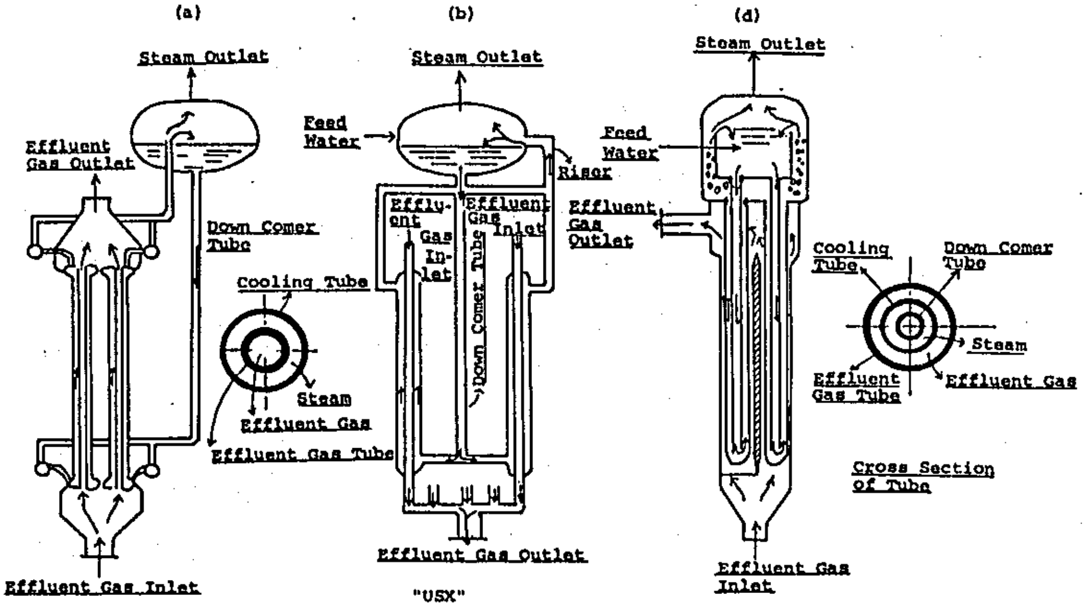
Developing Companies	Type	Characteristics
Lummus	SRT	Lummus first developed vertical crackers in which the thermal stress of pyrolysis coils and their supports were easily absorbed. This led to the first establishment of the high selectivity short residence time crackers. The design is intended to obtain high yield of olefin, i.e., ethylene and propylene, and to produce high reliability and economic performance by employing comparatively large diameter pyrolysis coils. The latest type, SRT-III, splits the coil into two parts in the radient zone.
Stone & Webster	USC	By employing comparatively small diameter pyrolysis coils, S&W is intending to realize the severest and shortest residence time operation for the highest yield of ethylene. Therefore, an ideal arrangement of the coils and burners is necessary to make the coils heat uniformly. This results in an increase of the volume of the furnaces. At present, this cracker yields the severest pyrolysis, has the records of operating heavy feedstock cracking, such as vacuum gas oil, especially by the pilot plants.
KTI formerly (Selas)		The past horizontal type cracker developed by Selas had the largest actual records of employment in the world. On the basis of this technical advantage in the engineering design of process heaters, KTI builds the most compact furnaces through the stringent marginal design in view of thermodynamic, and material durability. The branched pyrolysis coils at the inlet come to one coil in the furnace.

2-6-2 Quench System

The effluents of cracked gas at the outlets of cracking furnaces have temperatures up to 750°C to 900°C. The recovery of heat in the effluents is a vital factor for ethylene plant economics. Usually the heat is recovered by employing Transfer Line Heat Exchangers, which generate 100 atg to 140 atg high pressure steam having high utilization potential, and is used drive the turbines for the compression of cracked gas or refrigerant.

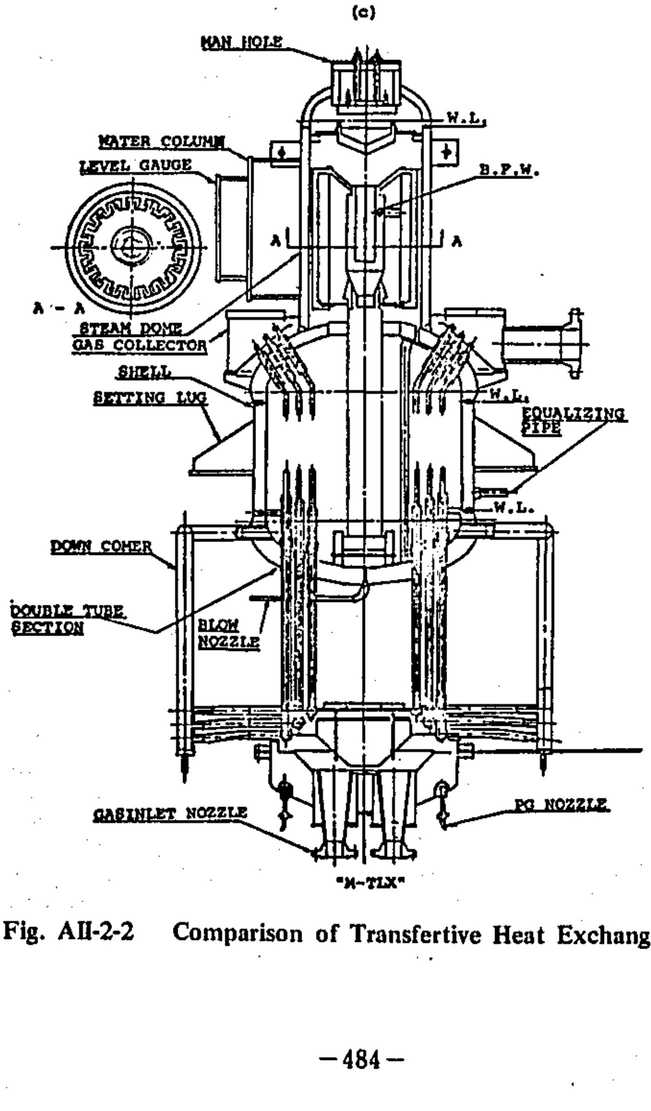
However, the formation of cokes and the generation of tars increases as the temperature in the TLE becomes lower because the heavier condensed fractions in the cracked gas polymerize. The cokes and tars stick to the walls of the heat exchanger tubes thereby increasing the pressure drop of the effluent in the TLE and decreasing the steam generation through detereoration of the heat transfer coefficient. Therefore, the allowable lowered temperature of the effluent at the outlet of the TLE is limited according to the properties of the cracked feedstocks and the characteristics of the TLE employed. This also leads to the decrease in generation of high pressure steam and accordingly to the lowering of thermal efficiency. Generally, the effluent can be lowered in the range of 400°C to 550°C, through the TLE, and the discharged effluent is rapidly cooled by mixing circulating quench oil and low pressure steam. In the pyrolysis of the vaccum gas oil by Exxon, reportedly the TLE can not be used because of severe formation of cokes and tars.

Therefore, the effluent gas must be quenched through directly contacting the circulating quench oils (the direct quench system). The heat in effluent gas is recovered at lower temperature, and is not utilizable as the high pressure steam. This leads to the lowering of the thermal efficiency in the plant operation. Furthermore, the highly viscous circulating quench oil increases and tends to polymerize to form cokes and tars at a lower temperature. Consequently, complexity will be increased in operation and design of the quench system. Representative types of TLE are shown in Figs. AII-2-1 and AII-2-2, and are outlined in Table AII-2-5.



"Schmidt" "USX" "Triple Tube Type"

Fig. AII-2-1 Comparison of Transfer Line Heat Exchangers (½)



"M-TLX"

Fig. AII-2-2 Comparison of Transfertive Heat Exchangers ($^2/_2$)

Table AII-2-5 The Comparison of Transfer Line Heat Exchangers

Item	Names & Developing Companies	Characteristics
a.	"Schmidt Boiler"	A multiple small sized double tube heat exchanger is used. This was the first TLE to effectively yield heat recovery through High Pressure Generation. The effluent gas flows downward inside the tube and the mixture of steam and saturated water rise in the shell side. A specific structure is applied to feed effluent from the pyrolysis coil to multiple tubes and to supply feed water to the multiple shells through single inlets. The small sized double tube gives a high heat transfer coefficient for the rapid quenching of effluent and for efficient steam generation.
b.	"USX" Stone & Webster	An ordinary type, comparatively large, double tube heat exchanger is used which installs a special protection device at the inlet of effluent. The effluent gas flows inside the tube and steam and water run in the shell side. Many USX's are installed on the side walls of furnaces connecting to each pyrolysis coil.
c.	"M-TLX" Mitsubishi Heavy Industry	By integrating the steam drum and multiple tubes of (a) type TLE, a compact TLE has been developed. The tubes in the steam drum are curved to absorb thermal stresses, due to the great difference of temperatures, between the inside and the outside. This leads to ease in on-line decoking which makes the decoking period significantly short (Ref. Fig. AII-2-2).
d.	Mitsui Petrochemical Industry and Mitsui Ship Building	As shown in Fig. AII-2-1, (d), the TLE consists of triple tubes; A downcomer of boiler feed water, a vaporizer of steam, and a shell in which effluent gas flows. As all of the tubes are hung downward from the end of the steam drum, the TLE is free from the any thermal stresses. This results in ease of on-line decoking. Because effluent gas passes through the shell side, the wall temperature of heat exchanging tube is higher, in the range of between 10 and several times ten degrees Centigrade, than that of other type TLE which results in preventing the condensation of effluent. Furthermore, the formed cokes on the outer side of tubes help the heat transfer by expanding the heat transfer areas. Therefore, the TLE permits the lower outlet temperature of effluent, and can obtain the favorable recovery of heat.

Furthermore, formation of cokes and tars increases the pressure drop not only in the transfer line heat exchanger tubes but also in the pyrolysis coils. Consequently, the pressure of the effluent in the coils will rise at the pyrolysis. The pressure rise in thermal cracking deteriorates the yields of olefins and their selectivity. Therefore, more frequent removal of adhering cokes and tars in the equipment i.e., decoking, is required when heavier feedstocks are used. This leads to the shorter furnace run period (decoking interval) and the complexity in operation.

2-7 The Employment of Gas Oil as Feedstocks in the Philippines

Fig. AII-2-3 shows the outline of effects of employing heavier feedstocks described previously, on the operation, and on the economy of an ethylene plant. The advantage of utilizing heavier petroleum fractions as a feedstock for pyrolysis is their low prices which is generally caused by their imbalance of supply and demand. However, as we have discussed already, the employment of heavier feedstocks for olefin production is not a matter of admitting hasty decision making. In comparison with the 90 to 120 days of decoking interval required for naphtha cracking, only 7 to 14 days, approximately 1/20 - 1/10 of that of naphtha cracking, are permitted in vaccum gas oil cracking. The drastically shortened furnace run period, as mentioned above, shows a rapid increase in technical difficulties as feedstocks become heavier. In addition to the difficulties due to the high sulfur content and lowered yields of olefins, the stringent tendency of the formation of cokes and tars in pyrolysis coils and transfer line heat exchangers causes serious problems in operations and deteriorates the efficiency of heat recovery from the effluent gas discharged from the pyrolysis coils.

Therefore, the pyrolysis of heavier feedstocks is attractive only when heavier petroleum fractions are available at sufficiently low prices. In the Philippines, the prices little differ from one another for the petroleum fractions which can be used as feedstocks, if the prices are evaluated on the refinery net back basis. Consequently, the lighter fractions are more preferable as far as the profit of the petrochemical complex is concerned.

To compensate for the lowered plant economy by employing heavier petroleum fractions, their prices must be sufficiently lowered. Reportedly, in gas oil cracking, the competitive price of gas oil is approximately 80% or lower of the price of naphtha even in view only of plant economy, although the further studies are necessary for final conclusion.

The employment of vaccum gas oil is not likely to be recommendable in view of stringent technical difficulties and the deterioration of plant economy due to the higher sulfur content and lower yields of olefins.

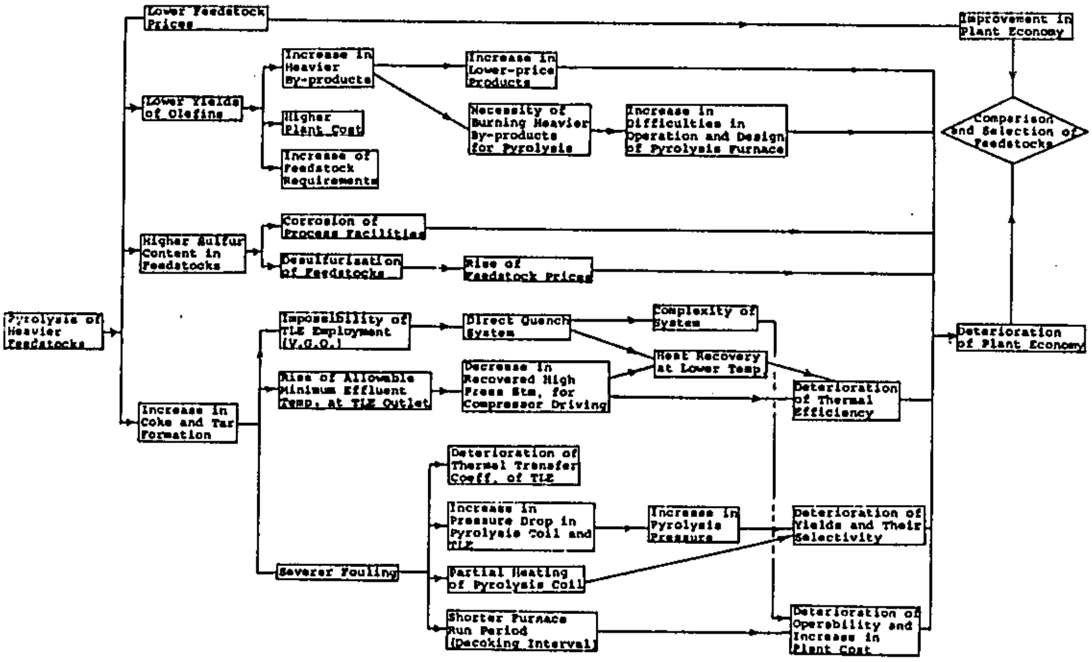


Fig. AII-2-3 Effects of Employing Heavier Feedstocks on Operation and Economy of an Ethylene Plant