IX. Raw Materials for Polyester (p-TPA, DMT)

1. Establishment of Production Capacity

1-1 Production Capacity in View of Demand Forecast

The major factor in deciding the plant capacity is the extent of demand for the product to be turned out. Detailed description has already been made in Chapter V concerning the demand forecast for raw materials for polyester covering the whole area of Southeast Asia and Indonesia. Table IX-1 shows excerpts of relative points therefrom.

Table IX-1 Indonesian Domestic Demand and Possible Export Amount to Southeast Asia of Raw Materials for Polyester

	Indonesian Domestic Demand (ton as TPA)	Possible Export to Southeast Asia (ton as TPA)
1974	23,000	-20,000
1975	33,000	42,000
1976	54,000	56,000
1977	80,000	166,000
1978	110,000	285,000
1979	140,000	472,000
1980	170,000	642,000
1981	182,000	775,000

As far as the years 1977 through 1978 are concerned, the forecast demand in Indonesia alone will be 95,000 t/y, and the exportable amount to Southeast Asian areas attain a level of approximately 230,000 tons. This signifies that the capacity for turning out 95,000 t/y will be necessary in order to fulfill the domestic demand. However, in the case of operating a so-called facilities-intensive industrial plants, the degree of expansion in production facilities must be carried out with a predesignated extent of scale. Therefore, in a number of cases, the design for the expansion is carried out firstly by forecasting the future demand increase and then by setting a certain period as a target in order to arrive at a production capacity design which exceeds the forecast level to a certain extent.

If a demand increase over forthcoming two years is taken into consideration on the basis of the above design policy, the suitable capacity level will be 160,000 t/y. Concerning the upper limit of the capacity it can be forecast that the export amount during the period from 1977 to 1978 will attain more than 200,000 t/y if the export promotion is effectively carried out. This signifies that a considerably large-scale plant construction is also possible.

However, in practice, the ex-factory price is often lower than the domestic market price due to the fact that the exportation has a higher extent of uncertain factors when compared with domestic sales, the existence of competition with other countries in international market and the necessity for incorporating into the price structure the factors such as transportation, import duties, trading company margin, etc. This being the circumstance, the incorporation of an excessive rate of exportation involves a high risk.

Therefore, it is recommendable that the rate of export allocation be controlled on a level less than 50%. In this sense, the practical maximum production capacity for the period from 1977 to 1978 will be 190,000 t/y.

1-2 Production Capacity in View of Economic Scale of a Plant

In a facilities intensive industry such as chemical industry, the production cost will become lower as the scale of facilities become larger due to the reduction in the extent of fixed-cost burden per unit amount of the product turned out. In order to construct a plant which has international competitiveness, it is necessary to adequately establish the optimum production scale in view of, and on the basis of, the worldwide trend of the market for the products.

Table V-36 shows the data regarding the new construction and expansion of raw material plant facilities for polyester now being projected in various countries of the world. According to the table, it seems that 60,000 t/y is the minimum economic scale so that the normal economic scale will be approximately 100,000 t/y. Some large-scale plants of more than 200,000 t/y production scale are also being projected.

The extent of the minimum economic scale is affected also by the amount of the works undertaken for the installation of the auxiliary facilities sectors which are to be simultaneously constructed with the main plant. In other words, in the case of constructing a plant in a site where the substantiation of not only the infrastructures but also all the off-site facilities are adequately achieved, the attainment of economic production operation can be amply undertaken even with a small-scale plant because of the low extent of the required facilities investment. On the other hand, when the amount of the works to be undertaken for the construction of the auxiliary facilities sectors is vast, the burden would be too great to be endured by a small-scale plant, thereby requiring in such an event a large-scale production plant.

In the case of Indonesia, it is considered imperative to place a particular emphasis on the industrial development in view of national governmental policies, so that the cases in this country would fall under the latter category. On the other hand, it should be taken into consideration in this connection that a further general increase in the scale of the plants has also been undertaken during the period from the compilation of Table V-36 up to the present. Therefore, in Indonesia, it seems necessary to set the minimum economic scale at 60,000 t/y or more, if possible, on a level of 100,000 t/y.

1-3 Limitations in View of Site Conditions

Once the plant site is finally selected, the problems pertaining to the limitation should be solved concerning such point as the securing of land usable for the construction, the size of the machinery and equipment which can be transported into the site, etc. However, in the case of Indonesia, as has been mentioned in Chapter VII, vast and flat lands are available in a number of suitable sites for chemical industrial construction. Also, no particular problems or impediments are present regarding the port facilities and roads for the transportation of the machinery and equipment up to the site. Therefore, it seems unnecessary to take into consideration the limitation factors in this respect.

However, care must be exercised regarding the p-xylene. For the purpose of securing stable supply and also for achieving saving in foreign currency it is necessary to consider as pre-requisite condition the domestic production of p-xylene. In other words, it is necessary to construct a raw material p-xylene production plant corresponding to the production capacity of the p-TPA/DMT plant. In case of adopting disproportionation of toluene, production of p-xylene will be 130, 000 to 160, 000 t/y. The amount of p-TPA/DMT to be produced from these p-xylene will be 210, 000 to 250, 000 t/y.

1-4. Conclusions Concerning the Production Capacities

As the conclusion drawn from the above scrutinizations, the following two cases will be taken up as the subjects for consideration.

Case 1: 150, 000 t/y

The production amount to fulfill the Indonesian domestic demand for the period from 1979 to 1980.

Case 2: 210, 000 t/y

The production amount corresponding to the increased production of p-xylene by adopting disproportionation of toluene (in this case the maximum exportation rate of p-TPA/DMT will be 50%).

2. Selection of Product Items

p-TPA and DMT are the two raw materials for polyester production and at present these two items are competing with each other. More than ten years in the past, the contentions have been undertaken concerning the comparative superiority between these two items and, therefore, it is difficult to forecast that the conclusion in this respect will be drawn in the foreseeable future. In view of the raw material production operation, the producers of raw materials are obliged to supply the types of raw materials which are required by the fiber plants. Therefore, the selection of the product items should be decided in accordance with the present as well as future demand. Nevertheless, it is necessary to pay full attention to the future trend of the demand in connection with the types of raw materials.

2-1 Historical Observation

When producing polyester, the easiest method is to directly react TPA with ethylene glycol. However, TPA does not melt even at a high temperature level of more than 300°C and also it is not soluble to most of the available solvents. This being the case, an economic refining process for TPA has long been sought after, however, there have been several technical problems such as the deterioration of polyester turned out because of the etherification of ethylene glycol at the time of esterification reaction. Due to this fact, a process was invented as one of the TPA refining methods in which the TPA is esterified into DMT which would be further refined, and then, interesterification with ethylene glycol is undertaken to produce polyester.

In 1953 when Du Pont industrialized the production of polyester for the first time in the world, DMT was employed as the raw material. Since then, efforts have been exerted in the technical improvements in the esterification process and refining process of TPA. As a result, several processes were invented for the production of p-TPA. Of these newly developed processes, the Amoco Process which was industrialized in 1965 was the most successfully developed process. Since then, along with the improvements made in the esterification technique, the production of p-TPA has drastically been increased so that ICI who thus far had been employing the conventional DMT process entirely switched over to the direct esterification process by employing p-TPA. This being the circumstance, the rate of p-TPA as the raw material for polyester has drastically been increased. Further detailed explanation on this point is made in Chapter V.

2-2 Observation in View of the Trend of New Construction or Expansion of Production Facilities

As has been described in detail in Chapter V, the new installation or expansion projects for the raw material plants for polyester show a trend that the extent of expansion for p-TPA production exceeds that for the DMT. Therefore, the future trend is forecast that the growth of TPA production will exceed that of DMT.

2-3 Observation in View of Demand Structure in Indonesia

At present, the companies which are now undertaking or projecting the production of polyester in Indonesia are Toray, Asahi Chemical, Teijin, Unitika, Kuraray, AKZO, Hoechst, etc. The decisions of these companies as to the selection of whether DMT or p-TPA as a raw material entirely depends on the economics and the level of technique of these companies as well as on the procurement availability of these raw materials. Globally, the rate of p-TPA as the raw material for polyester will increase in future and particularly, this trend is conspicuous in the new construction projects. Therefore, in Indonesia also, the number of companies who will produce polyester by employing p-TPA as a raw material will increase in the future.

However, because of the historical and technical background of the company, there may well be some who would insist in adopting the traditional DMT process. Therefore, it is difficult to confine the type of the selected raw material into one of the two at the time of carrying out case studies, thereby making it imperative to take up both DMT and p-TPA as the subjects for scrutinization of raw materials. Further, in the case of producing both of these raw materials simultaneously, it is reasonable to assume that the extent of production of DMT and p-TPA would be on an approximately the same level considering the future growth potential of p-TPA and the economic scale of production for each plants for producing these raw materials.

2-4 Comparison of Economic Viability

Economic viability problem is one of the major factors for forecasting as to whether p-TPA or DMT will assume the main trend in the future as the raw materials for polyester. In other words, the point pertains itself to the difference in the cost when producing fiber by utilizing each one of these two raw materials.

On the other hand, regarding the cost difference at the time of producing fiber by utilizing as the raw material either p-TPA or DMT has been described in Chapter V. As has been evident from the description, the advantage of p-TPA is evident. The reason for such an advantage on p-TPA is that when compared with the DMT, the unit consumption at the time of polymerization is less in the case of p-TPA, thereby reducing the methanol recovery cost.

2-5 Conclusions on the Selections of Product Items

It is assumed that the time for the construction of the plant concerned in Indonesia is 1977. By then, the rate comprised by p-TPA will be further grown globally. Therefore, if the emphasis is to be placed on the future potential and economics of the processes to be employed, it is natural that p-TPA should be selected as the main subject for consideration. However, in the case of studies to be undertaken in this writing, both p-TPA and DMT will be taken as the subjects for scrutinization in consideration of the past records of achievement made by DMT. Further, in the event of simultaneous production of both materials, the amount of p-TPA and DMT to be produced shall be assumed as being on an approximately the same level considering the future growth potential of p-TPA and the economic scale of production for both plants. Further, the assumption was made that the amount of polyester fiber produced by DMT and by p-TPA are approximately the same level. Therefore, taking the difference between the unit consumption of DMT and p-TPA for producing polyester fiber the size of DMT plant is determined to be 10% as large as that of p-TPA.

3. Selection of the Process

3-1 General Conditions for the Selection of Process

The conditions affecting the decision concerning the selection of the process boils down to the point whether a certain process is superior to the others in economic view. However, it is also necessary to take into consideration the following factors which do not easily reflect their effects on the normally undertaken economic calculations.

(1) The Process must have Sufficient Records in Commercial Production

A high level of technological development ability, time and funds will be necessary in order to complete a process into industrialization. Therefore, if seemingly attractive processes are on a stage of development at which no sufficient allowance for these factors is available, the selection should be made from other processes which have amply proved their production capacity through already established industrial operation records.

(2) The Process must have Future Development Potential

No matter how brilliant the actual records of achievement may be, the introduction should be avoided if such a process has already become obsolete in the developed countries. Therefore; at the time of introducing a technique for the first time, the selection must be made from those which at that time is the latest and most advanced, thereby allowing little chance of immediate replacement by better processes.

(3) The Process must be Stable

In many cases of developing countries, the substantiation in the related industries, educational institutions, etc. which supports the operation of chemical industrial plants are not sufficiently achieved. Therefore, there are frequent cases in which the procurement of repair or replacement parts is difficult. Therefore, even if the economics of the process is slightly inferior, the process to be selected must be of such that the operation is not excessively sensitive to the variation in operational conditions and should be free from unique devices.

(4) The Products Turned out from the Selected Process should have a Vast Market

In the case of chemical industry, advantage would be greater if a plant scale is made as large as possible in order to reduce the burden of the fixed cost. The products will not only be supplied to the domestically operating fiber manufacturers but will also have to be exported in some portion.

Therefore, the quality of the product to be turned out should not only satisfy the utilization quality standards of specific fiber companies, but should be of such that will be internationally acceptable. In the case of producing products which are internationally applicable, the advantage is also obvious in view of the fact that the domestic demands can be fulfilled by means of importation from overseas during the period up to the completion of expansion of the plant facilities in the event that the domestic demand has grown higher than the existing plant capacity.

(5) The Process must be of the Resource-Saving and Energy-Saving Characteristics.

Along with the advancement in the reflection upon the thus-far conducted wasteful utilization of natural resources, it is evidently forecast that the price of the resources will further increase. Therefore, the factor of resource saving will become one of the most important criteria for the evaluation of future chemical industrial process.

3-2 Raw Material Routes and Polyester Production

There are several available routes for conducting polyester production. Figure IX-1 and Table IX-2 show a summary of the main routes in this connection. When broadly classified, the routes can be set into two categories, i.e., the one which goes through DMT and another, p-TPA. Further, the processes for producing DMT and p-TPA will be classified into several categories. For the most part, the starting raw material is p-xylene at the present stage. In the past, phthalic acid anhydride was employed; however, the raw material production for polyester by means of this process has entirely been discarded. Aside from p-xylene, toluene (or benzoic acid) are being used in an extremely limited sphere. In the following paragraphs, the production process for the various raw materials for polyester mentioned above will be described in accordance with the illustration made in Figure IX-1 and Table IX-2.

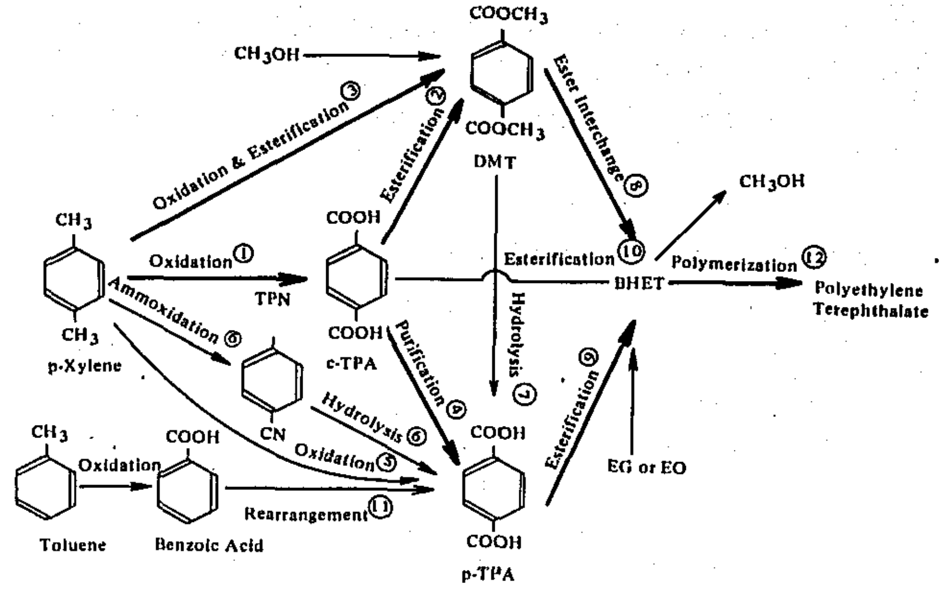


Figure IX-1 Various Processes of Monomer of Polyester and Polyester Production

(1) Oxidation of p-Xylene

This is a process for producing c-TPA by oxidizing p-xylene. The processes available in this field are Amoco Process, Mobil Process, Eastman Kodak Process, Toray Process, Nitric Acid Oxidation Process, etc. Of these, the most popularly employed is the Amoco Process in which bromine compounds are employed as the reaction accelerating agents.

Although the nitric acid oxidation process was popularly employed in the past, this process is almost entirely discarded at present. In the other processes, ketone or aldehyde is employed as the reaction accelerating agent.

(2) Esterification of c-TPA

This is a process for the production of DMT by esterification of c-TPA with methanol. Under the existence of excess amount of methanol, the esterification is carried out at high temperature and high pressure. The refining of DMT is carried out by the combination of the recrystallization process, chemical treatment, fractionation, etc. There are a number of esterification processes for producing c-TPA and the technology in this field is highly popularized.

Table IX-2 : Monomer of Polyester and Various Polyester Production Processes

1. Oxidation of p-Xylene
    - 1.1 Nitric Acid Oxidation
    - 1.2 Mid-Century or Amoco Process
        - Air oxidation promoted by bromine compounds
    - 1.3 Olin Mathieson or Mobil Process
        - Air oxidation promoted by methyl ethyl ketone
    - 1.4 Eastman Kodak Process
        - Air oxidation promoted by acetaldehyde
    - 1.5 Toray Process
        - Air oxidation promoted by paraldehyde
2. Esterification of c-TPA with Methanol followed by Purification
3. Witten Process
    - Successive oxidation and esterification via methyl p-methylbenzoate
4. Purification of c-TPA
    - 4.1 Amoco Process
        - Hydrogenation in water solution
    - 4.2 Mobil Process
        - Sublimation in steam
    - 4.3 Mitsubishi Process
    - 4.4 Toray Process
5. Maruzen Process
    - Modified air oxidation promoted by bromine compounds
6. Lummus Process
    - Ammoxidation of p-Xylene
    - Hydrolysis of Terephthalonitrile
7. Hercules Process
    - Hydrolysis of DMT with water
8. Ester Interchange of DMT with Ethylene Glycol
9. Esterification of p-TPA with Ethylene Glycol
10. Esterification of c-TPA with Ethylene Oxide
11. Rearrangement of Benzoic Acid
12. Polymerization of Bis-β-Hydroxyethyl Terephthalate

(3) Witten Process

This is a process for producing DMT without going through TPA by successive oxidation and esterification of p-xylene.

This process is otherwise known as Hercules Process.

(4) Refining of c-TPA

This is a process for producing p-TPA by refining c-TPA. Available processes are Amoco, Mobil, Mitsubishi and Toray processes. In the Amoco Process, c-TPA is made into a slurry by adding water and under high temperature and high pressure it is treated with hydrogen. The Mobil Process undertakes the sublimation refining, however, Mobil gave up the production of p-TPA through this process in 1973.

(5) Production of p-TPA from p-Xylene

This is a process for producing p-TPA simply by the oxidation reaction of p-xylene by means of an improved process of Amoco. This process has been industrialized in Matsuyama Petrochemical Industry Co.

(6) Lummus Process

This is a process to produce p-TPA by ammoxidation of p-xylene. In this process, p-xylene is oxidized by ammonia to terephthalonitrile and further terephthalonitrile is hydrolysed to p-TPA. No commercial plant has been built employing this process.

(7) Production of p-TPA from DMT

This is a process for producing p-TPA by means of hydrolysis of DMT. Commercial plant using this process has been built by Hercules, however its commercial operation has not been announced. When the production of DMT is undertaken by means of the Witten Process, it is not possible to produce p-TPA by refining c-TPA due to the fact that the process does not go through TPA. Therefore, the production of p-TPA is undertaken by carrying out the hydrolysis of DMT. It must be noted that question remains concerning the economics of this process.

(8) Interesterification of DMT and Ethylene Glycol

By means of the interesterification of DMT and ethylene glycol, the production of BHET (Bis-β-Hydroxyethyl-Terephthalate) which is the oligomer of polyester. At this reaction, methanol is removed.

(9) Esterification of p-TPA

By means of the esterification reaction of p-TPA and ethylene glycol, the production of BHET is carried out. This is otherwise known as the direct esterification process.

(10) Reaction between c-TPA and Ethylene Oxide

By adding ethylene oxide to c-TPA, BHET is produced. The produced BHET is refined and then polymerized. Although this process was carefully studied by several companies, no commercial operation has so far been achieved.

(11) Henkel Process

This is a process to produce p-TPA from toluene. Benzoic acid is produced by oxidation of toluene. From 2-mol of potassium benzoate, 1-mol each of dipotassium terephthalate and benzene is obtained by Henkel rearrangement reaction. Then, dipotassium terephthalate is hydrolyzed to p-TPA. This process has been industrialized by Mitsubishi Chemical Ind.

3-3 Selection of the TPA Process

Figure IX-1 shows the raw material production processes for polyester which have already been industrialized in the world at present including the DMT production. As the raw materials for TPA/DMT, the available items are p-xylene, toluene and phthalic acid anhydride, however, at present p-xylene takes up the most part of the raw materials. Regarding p-TPA, most of the presently marketed products consist of the one produced by the Amoco Process. The Mobil Process used to be employed; however, the plant was closed in 1973 because of the impossibility in competing with the Amoco Process. Also, it is reported that Hercules completed a plant for the first time in the world for the production of p-TPA by means of the hydrolysis of DMT. However, no final evaluation has been made in the economic and quality considerations. At any rate, no economic advantage can be anticipated in the DMT hydrolysis process.

In addition to the above, a number of processes known as p-TPA production process are being operated; however, all of these are the specific processes employed for the production of raw material plants owned by the fiber manufacturers themselves or by their subsidiaries and therefore, they are not to be within the scope of selection for a process to produce products to be distributed to a plural number of fiber companies or their subsidiaries. The most popularly employed process in the world today for producing p-TPA is the Amoco Process.

Table IX-3 shows the unit consumption of the Amoco Process.

Table IX-3 Unit Consumption of Amoco Process p-TPA Production

	Unit Consumption (/kg p-TPA)
Raw Materials	
p-Xylene	0.69 kg
Acetic Acid	0.096 "
Catalysts & Chemicals	0.61 US¢
Utility	
Fuel	0.80 kg

3-4 Selection of DMT Processes

Unlike TPA, DMT does not present conspicuous differences in the quality amongst the producers so that there would be no problem in the marketing aspect no matter which process is selected. Also, in view of the fact that DMT production has been undertaken for a long period of time in the past, it can be considered that the processes are stable technically; however, no great expectations can be had concerning the future potential in such aspects as the process improvement or innovation. Therefore, the problem to be considered here is the economy of the process.

As the typical processes for DMT production, the selection is made for the Witten Process which does not go through TPA and the Amoco Process which goes through TPA. Table IX-4, 5 show the unit consumption of these processes.

Table IX-4 Unit Consumption of Witten Process DMT Production

	Unit Consumption (/kg DMT)
Raw Materials	
p-Xylene	0.66 kg
Methanol	0.41 "
Catalysts & Chemicals	0.28 US¢
Utility	
Fuel	1.62 kg

Table IX-5 Unit Consumption of DMT Production via c-TPA

	Unit Consumption (/kg DMT)
Raw Materials	
p-Xylene	0.58 kg
Acetic Acid	0.07 "
Methanol	0.37 "
Catalysts & Chemicals	0.28 US¢
Utility	
Fuel	0.71 kg

3-5 Process for Simultaneous Production of p-TPA and DMT

When the production of both p-TPA and DMT is to be considered as the pre-requisite condition, the following should be taken into consideration.

In addition to the already described processes for separate production of p-TPA and DMT, there is a process in which production of DMT and p-TPA can be undertaken from c-TPA. In this case, the economic advantages are apparent due to the possibility of constructing a large-scale plant for the production of the raw material c-TPA to be fed to both the p-TPA and DMT plants. Further, the combination of p-TPA and DMT plants in such a manner that the control of the production in accordance with the demand trend is attractive not only in view of the cost consideration but also of the flexibility against the demand fluctuation. Table IX-6 shows the unit consumption of the process.

Table IX-6 Unit Consumption of p-TPA and DMT Production via c-TPA (Ratio of p-TPA and DMT, 1:1.1)

	Unit Consumption (/kg)
Raw Materials	
p-Xylene	0.632 kg
Acetic Acid	0.0824 "
Methanol	0.194 "
Catalysts & Chemicals	0.437 US¢
Utility	
Fuel	0.753 kg

3-6 Conclusion of the Process Selection

By summarizing the above scrutinizations, the selection is made for Amoco Process for the production of p-TPA and Witten Process for DMT. The Amoco Process is also selected for the simultaneous production of p-TPA and DMT via c-TPA.

4. Process Description

4-1 Amoco Process: TPA and DMT Processes

Amoco Process can be separated into two parts, i.e., one for the production of c-TPA by the oxidation of p-xylene and another for undertaking c-TPA refining in order to produce p-TPA or DMT by esterification. Figure IX-2 shows the flowsheet of these processes.

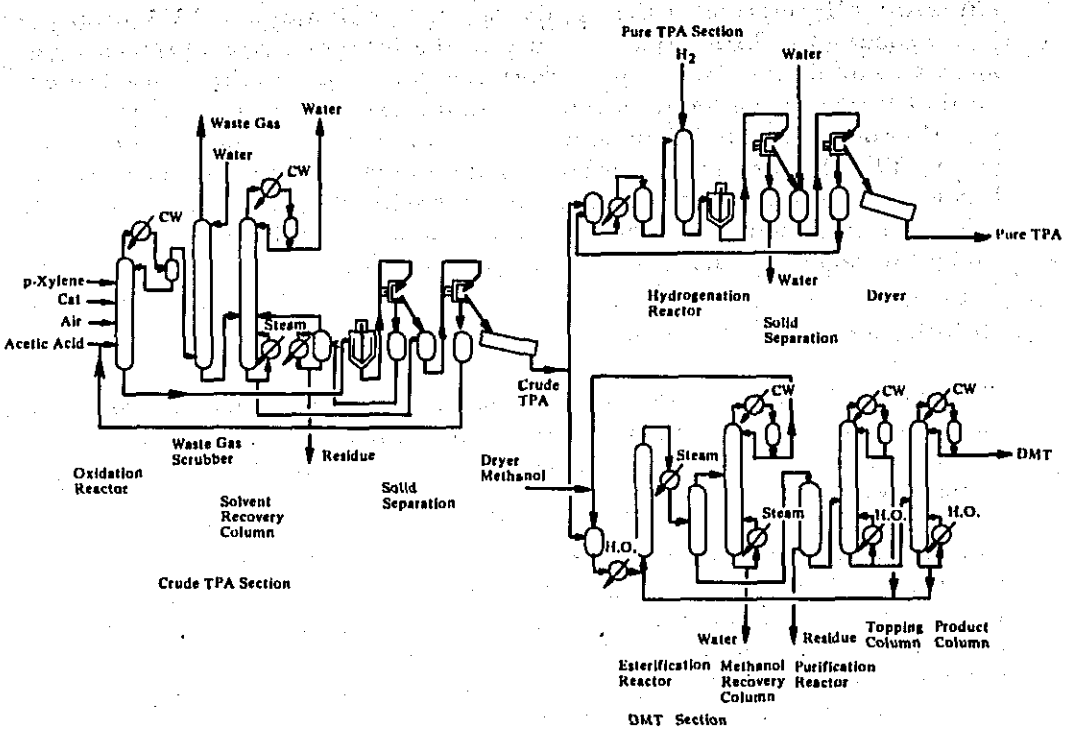
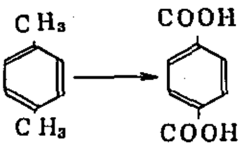


Figure IX-2 Flow Diagram of Amoco TPA/DMT Process

(1) Production of c-TPA

The Amoco's c-TPA process was developed in 1955 by A. Saffer, et. al. of Mid Century, and was industrialized for the first time in the world in 1958 by Mitsui Petrochemical. A number of improvements have since been incorporated and the completed process of today has been achieved.

In this process, the TPA is obtained by one stage from p-xylene in which the oxidation of two methyl-radicals is carried out by means of a strong hydrogen abstracting effect displayed by the addition of bromine compounds.



In this process, p-xylene is fed to the oxidation reactor together with the solvent acetic acid in which it is oxidized up to TPA by 1-pass. The reaction conditions are; temperature 200°C approximately, pressure 10 to 30 kg/cm^2 Hg, reaction time within one hour. As the catalysts, cobalt, manganese and other heavy metals and the above-mentioned bromine compounds are employed.

The produced TPA is suspended in the solvent or the slurry status which is separated by means of a centrifugal separator and further, in order to eliminate catalysts, etc., the washing by acetic acid is carried out. After the elimination of the catalysts, it is dried and c-TPA is produced. The produced c-TPA further contains the intermediate reaction compounds such as aldehydes, e.g., p-formyl benzoic acid or other tinted substances so that it cannot be used as the raw materials for polyester as it is. Developments are being undertaken to obtain a process to produce high purity TPA by means of controlling the oxidation reaction condition. However, no popularization has so far been achieved. In the normal cases, the refining is being undertaken by the under-mentioned refining process or by the DMT process.

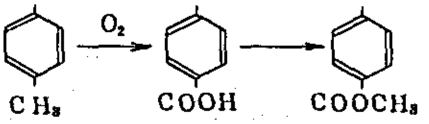
(2) Refining of c-TPA

This is a process developed by Amoco in 1965 as a process for eliminating from c-TPA the aldehydes and the tinted impurities. The water solution of TPA is brought into contact with hydrogen in order to reduce the undesired impurities.

c-TPA is dissolved into the water, the amount of which is several times the volume of p-TPA, at a high temperature and then will be fed to the reduction reactor together with hydrogen. Precious metal catalysts are filled inside the reactor by employing carbon as the carrier and the undesired impurities are reduced by hydrogen and then eliminated. The reacted substance is cooled and the TPA is recovered in the form of crystal. Separation is carried out by means of centrifugal separation and washing will then be carried out in order to further eliminate the remaining impurities. The production of p-TPA is then undertaken by the separation and drying.

(3) Production of DMT

As and when necessary, c-TPA can be refined into DMT. c-TPA is esterified by means of methanol to be turned into crude DMT. The crude DMT has conventionally been refined by the crystallization process, however, in the recent years, a number of processes are employing the refining method by means of chemical reaction and fractionation instead of crystallization for the purpose of cost reduction. In this refining method, the elimination of the aldehydes and the tinted impurities is undertaken by the chemical reactions and fractionation; due to the fact that fractionation alone cannot sufficiently carry out the elimination. Eventually, the refined DMT is produced by fractionation.



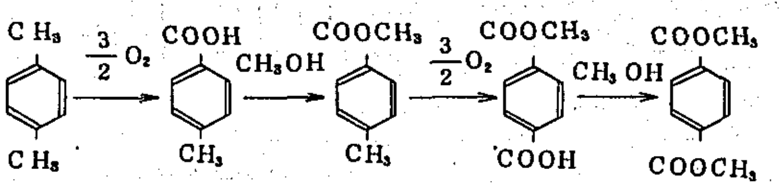
The feature of the Amoco Process is the employment of such corrosive and specific substances as acetic acid and bromine compounds in order to proceed the reaction. Due to this feature, a process can be made extremely simple and the facilities itself can be designed highly compact. As a result, a high extent of yield and a low level of utility consumption have been achieved.

However, in view of the material requirements for composing the equipment, the necessity for abundant utilization of titanium which is an anti-corrosive material invites some extent of additional cost in constructing the reaction equipment. In spite of this disadvantage, the consumption of the raw materials and utilities will be reduced, thereby achieving a low level of fixed cost and variable cost.

4-2 Witten DMT Process

The Witten DMT Process was invented in 1950 by I.E. Levine of C.R.C. and in 1951 by E. Katzschmanu of Imhausen. The industrialization of this process was achieved by Hercules for the first time and then followed by Imhausen and Hoechst.

This process produces DMT by means of four stageous of oxidation and esterification reactions from p-xylene. The reactions will progress as shown below.



Generally speaking, the oxidation of the first methyl-radical is easy and the oxidation of the second methyl-radical is difficult. However, if the esterification of the already oxidized carboxyl-radical is undertaken, the oxidation of the second methyl-radical becomes easy.

The Witten Process takes the advantage of this phenomenon so that the oxidation of the two methyl-radicals of the p-xylene is undertaken. Figure IX-3 shows the flowsheet of the Witten Process. p-Xylene is fed to the oxidation reactor together with the catalyst in which the air oxidation is carried out to produce p-toluic acid. In the same reactor, the oxidation of methyl p-toluate is also carried out simultaneously, thereby producing mono-methylterephthalate. The reaction conditions are; at 150°C, pressure 6 kg/cm^2, reaction time for ten hours. As the catalysts, salts of such heavy metals as cobalt, manganese, etc., are employed. No special solvent is used in this process. Adequately controlled p-xylene serves as a function of solvents in the reaction.

The oxidation reaction compounds then is fed to the esterification reactor in which the production of crude DMT and methyl p-toluate is carried out. The esterification reaction compounds are then separated into DMT and methyl p-toluate by means of fractionationand the former is fed to the refining process and the latter recycled to the oxidation reactor.

The refining of crude DMT is carried out by recrystallization and fractionation on two stages. In other words, the crude DMT is dissolved in methanol and then crystallized by means of cooling. Then, it is separated from the mother liquor by means of a centrifugal separator. Due to the fact that no sufficient purity is obtainable by 1-stage crystallization, the identical operation is repeated. DMT coming out from the two-stage crystallization is then refined by fractionation in order to produce refined DMT.

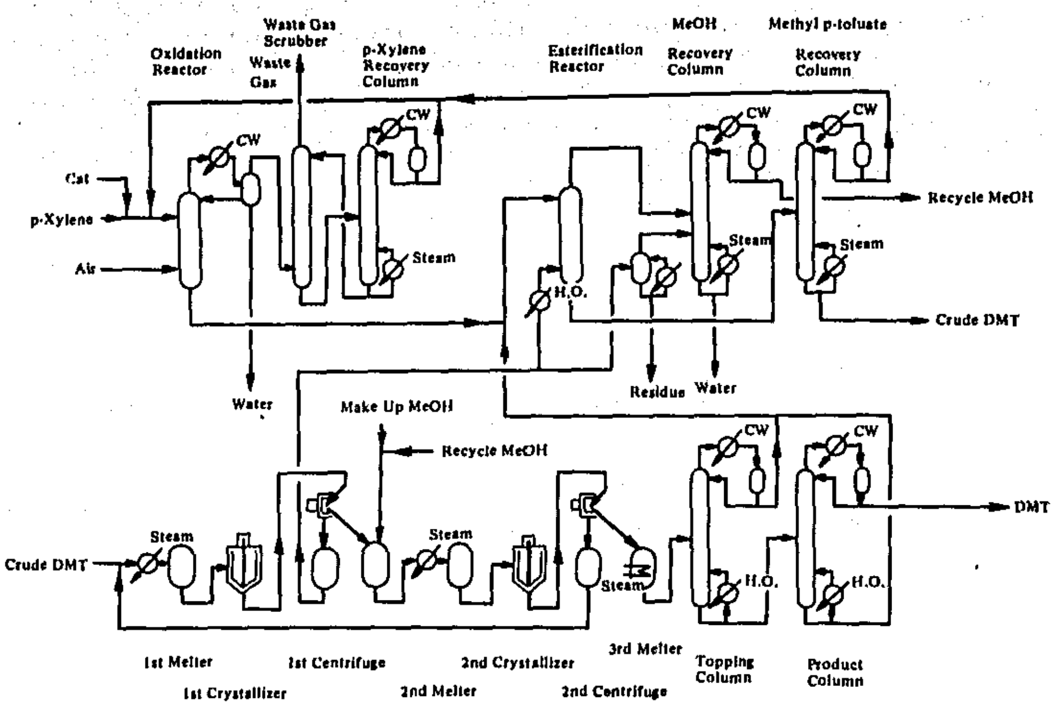


Figure IX-3 Flow Diagram of Witten Process

The feature of the Witten Process is that the reaction goes through ester in order to oxidize the second methyl-radical, the oxidation of which is highly difficult under the normal reaction conditions. Because of this feature, no such corrosive substances such as acetic acid or bromine compounds is involved in this reaction, thereby necessitating no special anti-corrosive materials in the process equipment structure. Also due to the fact that DMT is directly obtainable from p-xylene, it is possible to integrate the process into a single plant, thereby reducing the extent of construction cost for reaction facilities.

On the other hand, the conversion rate in the oxidation reaction is low and the extent of various impurities produced by sub-reactions is high, so that the consumption of the raw material is also high. Because of the necessity for carrying out a large amount of recycling and also due to the necessity of employing the crystallization process for refining, the extent of utility consumption will become high. As a result, the variable cost and fixed cost of the utilities will be higher when compared with the other processes.

5. Construction and Operation of Plants

5-1 Construction Schedule

Figure IX-4 shows a tentative schedule with a target of completion of the plant construction by mid 1977 when the test operation commencement will be undertaken.

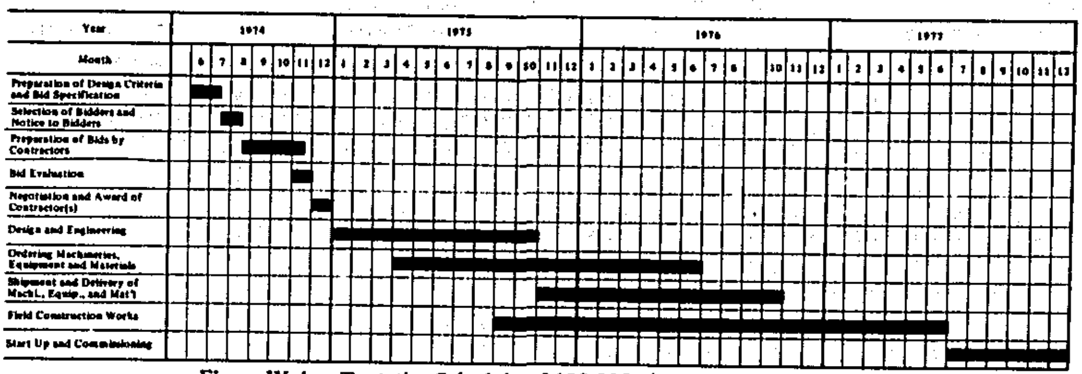


Figure IX-4 Tentative Schedule of 100,000 t/y p-TPA Plant Construction

5-2 Estimate on the Construction Cost

A detailed explanation was already made in the Chapter pertaining to the Method of Survey regarding the scope and contents of the construction cost estimate.

Tables IX-7 through IX-15 show the investment breakdown required for the construction of p-TPA, DMT plant in Indonesia for target of a completion by mid 1977. The total investment is US$93 million for the construction of a p-TPA plant for the capacity of 100,000 t/y and, US$98 million for a Witten DMT process plant of 110,000 t/y. When constructing a plant for the production of 50,000 t/y p-TPA and 55,000 t/y of DMT via c-TPA, the estimated amount will be US$100 million.

Table IX-7 Estimated Capital Requirement (1977) Case A-1

Case A-1 p-TPA 50,000 t/y
DMT 55,000 t/y

(Base Year 1971)

	Rupiah (10^6)			U.S.$ Equiv. (10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchange Portion	Local Currency Portion	Total
1. Equipment						
Plant	5,135	-	5,135	12,375	-	12,375
Utilities Plant & Facilities	1,975	-	1,975	4,758	-	4,758
Auxiliary & Off-Site	707	-	707	1,703	-	1,703
Sub-Total	7,817	-	7,817	18,836	-	18,836
2. Construction Costs						
Installation	1,004	559	1,563	2,419	1,347	3,766
Civil Works	807	1,009	1,816	1,944	2,431	4,375
Office Bldg. & Housing Colony	258	803	1,061	622	1,936	2,558
Sub-Total	2,069	2,371	4,440	4,985	5,714	10,699
3. Transportation	2,111	-	2,111	5,086	-	5,086
4. Plant Cost	11,997	2,371	14,368	28,907	5,714	34,621
5. Escalation & Contingency	8,185	1,617	9,802	19,722	3,897	23,619
6. Total Plant Costs	20,182	3,988	24,170	48,629	9,611	58,240
(1. Process Plant)	12,661	940	13,601	30,507	2,267	32,774
(2. Auxiliary & Off-Site)	7,086	1,696	8,782	17,075	4,086	21,161
(3. Buildings)	435	1,352	1,787	1,047	3,258	4,305
7. Royalty, Eng. Fee, Technical Exp.	5,823	-	5,823	14,300	-	14,300
8. Pre-Operating Expenses	-	897	897	-	2,162	2,162
9. Land	-	671	671	-	1,617	1,617
10. Interest During Construction	2,071	-	2,071	4,990	-	4,990
11. Working Capital	-	7,757	7,757	-	18,693	18,693
12. Total Capital Requirement	28,076	13,313	41,389	67,919	32,083	100,002

Table IX-8 Estimated Capital Requirement (1977) Case A-2

Case A-2 p-TPA 70,000 t/y
DMT 77,000 t/y

(Base Year 1971)

	Rupiah (10^6)			U.S.$ Equiv. (10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchange Portion	Local Currency Portion	Total
1. Equipment						
Plant	6,468	-	6,468	15,586	-	15,586
Utilities Plant & Facilities	2,499	-	2,499	6,022	-	6,022
Auxiliary & Off-Site	855	-	855	2,061	-	2,061
Sub-Total	9,822	-	9,822	23,669	-	23,669
2. Construction Costs						
Installation	1,222	705	1,927	2,944	1,700	4,644
Civil Works	1,019	1,248	2,267	2,456	3,008	5,464
Office Bldg. & Housing Colony	311	858	1,169	750	2,067	2,817
Sub-Total	2,552	2,811	5,363	6,150	6,775	12,925
3. Transportation	2,653	-	2,653	6,392	-	6,392
4. Plant Cost	15,027	2,811	17,838	36,211	6,775	42,986
5. Escalation & Contingency	10,252	1,918	12,170	24,703	4,622	29,325
6. Total Plant Costs	25,279	4,729	30,008	60,914	11,397	72,311
(1. Process Plant)	15,875	1,187	17,062	38,253	2,861	41,114
(2. Auxiliary & Off-Site)	8,881	2,100	10,981	21,400	5,061	26,461
(3. Buildings)	523	1,442	1,965	1,261	3,475	4,736
7. Royalty, Eng. Fee, Technical Exp.	8,058	-	8,058	19,415	-	19,415
8. Pre-Operating Expenses	-	1,256	1,256	-	3,026	3,026
9. Land	-	792	792	-	1,908	1,908
10. Interest During Construction	2,632	-	2,632	6,343	-	6,343
11. Working Capital	-	10,607	10,607	-	25,558	25,558
12. Total Capital Requirement	35,969	17,384	53,353	86,672	41,889	128,561

Table IX-9 Estimated Capital Requirement (1977) Case A-3

Case A-3 p-TPA 100,000 t/y
DMT 110,000 t/y

(Base Year 1971)

	Rupiah (10^6)			U.S.$ Equiv. (10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchang Portion	Local Currency Portion	Total
1. Equipment						
Plant	8,302	-	8,302	20,002	-	20,002
Utilities Plant & Facilities	3,208	-	3,208	7,731	-	7,731
Auxiliary & Off-Site	1,060	-	1,060	2,556	-	2,556
Sub-Total	12,570	-	12,570	30,289	-	30,289
2. Construction Costs						
Installation	1,518	903	2,421	3,658	2,175	5,833
Civil Works	1,318	1,585	2,903	3,175	3,819	6,994
Office Bldg. & Housing Colony	393	914	1,307	947	2,203	3,150
Sub-Total	3,229	3,402	6,631	7,780	8,197	15,977
3. Transportation	3,394	-	3,394	8,178	-	8,178
4. Plant Cost	19,193	3,402	22,595	46,247	8,197	54,444
5. Escalation & Contingency	13,093	2,321	15,414	31,550	5,592	37,142
6. Total Plant Costs	32,285	5,722	38,007	77,797	13,789	91,586
(1. Process Plant)	20,287	1,518	21,805	48,886	3,658	52,544
(2. Auxiliary & Off-Site)	11,336	2,666	14,002	27,317	6,425	33,742
(3. Buildings)	662	1,538	2,200	1,594	3,706	5,300
7. Royalty, Eng. Fee, Technical Exp.	11,199	-	11,199	26,986	-	26,986
8. Pre-Operating Expenses	-	1,794	1,794	-	4,323	4,323
9. Land	-	947	947	-	2,283	2,283
10. Interest During Construction	3,409	-	3,409	8,215	-	8,215
11. Working Capital	-	14,829	14,829	-	35,733	35,733
12. Total Capital Requirement	46,893	23,292	70,185	112,998	56,128	169,126

Table IX-10 Estimated Capital Requirement (1977) Case B-1

Case B-1 p-TPA 70,000 t/y

(Base Year 1971)

	Rupiah (10^6)			U.S.$ Equiv. (10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchange Portion	Local Currency Portion	Total
1. Equipment						
Plant	3,802	-	3,802	9,162	-	9,162
Utilities Plant & Facilities	1,653	-	1,653	3,983	-	3,983
Auxiliary & Off-Site	567	-	567	1,367	-	1,367
Sub-Total	6,022	-	6,022	14,512	-	14,512
2. Construction Costs						
Installation	784	431	1,215	1,889	1,039	2,928
Civil Works	621	783	1,404	1,497	1,886	3,383
Office Bldg. & Housing Colony	175	662	837	422	1,594	2,016
Sub-Total	1,580	1,876	3,456	3,808	4,519	8,327
3. Transportation	1,625	-	1,625	3,917	-	3,917
4. Plant Cost	9,227	1,876	11,103	22,237	4,519	26,756
5. Escalation & Contingency	6,295	1,280	7,575	15,169	3,083	18,252
6. Total Plant Costs	15,522	3,156	18,678	37,406	7,602	45,008
(1. Process Plant)	9,438	726	10,164	22,745	1,747	24,492
(2. Auxiliary & Off-Site)	5,789	1,316	7,105	13,950	3,172	17,122
(3. Buildings)	295	1,114	1,409	711	2,683	3,394
7. Royalty, Eng. Fee, Technical Exp.	4,231	-	4,231	10,195	-	10,195
8. Pre-Operating Expenses	-	592	592	-	1,426	1,426
9. Land	-	512	512	-	1,233	1,233
10. Interest During Construction	1,576	-	1,576	3,797	-	3,797
11. Working Capital	-	5,312	5,312	-	12,800	12,800
12. Total Capital Requirement	21,329	9,572	30,901	51,398	23,061	74,459

Table IX-11 Estimated Capital Requirement (1977) Case B-2

Case B-2 p-TPA 100,000 t/y

(Base Year 1971)

	Rupiah (10^6)			U.S.$ Equiv. (10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchange Portion	Local Currency Portion	Total
1. Equipment						
Plant	4,542	-	4,542	10,945	-	10,945
Utilities Plant & Facilities	1,975	-	1,975	4,758	-	4,758
Auxiliary & Off-Site	656	-	656	1,580	-	1,580
Sub-Total	7,173	-	7,173	17,283	-	17,283
2. Construction Costs						
Installation	907	513	1,420	2,187	1,236	3,423
Civil Works	739	918	1,657	1,780	2,211	3,991
Office Bldg. & Housing Colony	203	688	891	489	1,658	2,147
Sub-Total	1,849	2,119	3,968	4,456	5,105	9,561
3. Transportation	1,937	-	1,937	4,667	-	4,667
4. Plant Cost	10,959	2,119	13,078	26,406	5,105	31,511
5. Escalation & Contingency	7,476	1,446	8,922	18,014	3,483	21,497
6. Total Plant Costs	18,435	3,565	22,000	44,420	8,588	53,008
(1. Process Plant)	11,230	864	12,094	27,059	2,080	29,139
(2. Auxiliary & Off-Site)	6,864	1,544	8,408	16,539	3,719	20,258
(3. Buildings)	341	1,157	1,498	822	2,789	3,611
7. Royalty, Eng. Fee, Technical Exp.	5,764	-	5,764	13,889	-	13,889
8. Pre-Operating Expenses	-	846	846	-	2,038	2,038
9. Land	-	609	609	-	1,467	1,467
10. Interest During Construction	1,917	-	1,917	4,620	-	4,620
11. Working Capital	-	7,285	7,285	-	17,555	17,555
12. Total Capital Requirement	26,116	12,305	38,421	62,929	29,648	92,577

Table IX-12 Estimated Capital Requirement (1977) Case C-1

Case C-1 DMT 77,000 t/y

(Base Year 1971)

	Rupiah (10^6)			U.S.$ Equiv.(10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchange Portion	Local Currency Portion	Total
1. Equipment						
Plant	3,485	-	3,485	8,397	-	8,397
Utilities Plant & Facilities	2,345	-	2,345	5,650	-	5,650
Auxiliary & Off-Site	564	-	564	1,358	-	1,358
Sub-Total	6,394	-	6,394	15,405	-	15,405
2. Construction Costs						
Installation	810	458	1,268	1,953	1,103	3,056
Civil Works	797	960	1,757	1,919	2,314	4,233
Office Bldg. & Housing Colony	248	714	962	597	1,719	2,316
Sub-Total	1,855	2,132	3,987	4,469	5,136	9,605
3. Transportation	1,726	-	1,726	4,158	-	4,158
4. Plant Cost	9,975	2,132	12,107	24,032	5,136	29,168
5. Escalation & Contingency	6,805	1,454	8,259	16,397	3,503	19,900
6. Total Plant Costs	16,780	3,586	20,366	40,429	8,639	49,068
(1. Process Plant)	8,809	770	9,579	21,220	1,856	23,076
(2. Auxiliary & Off-Site)	7,554	1,616	9,170	18,203	3,891	22,094
(3. Buildings)	417	1,200	1,617	1,006	2,892	3,898
7. Royalty, Eng. Fee, Technical Exp.	3,209	-	3,209	7,733	-	7,733
8. Pre-Operating Expenses	-	804	804	-	1,939	1,939
9. Land	-	546	546	-	1,317	1,317
10. Interest During Construction	1,635	-	1,635	3,941	-	3,941
11. Working Capital	-	6,724	6,724	-	16,202	16,202
12. Total Capital Requirement	21,624	11,660	33,284	52,103	28,097	80,200

Table IX-13 Estimated Capital Requirement (1977) Case C-2

Case C-2 DMT(W) 110,000 t/y

(Base Year 1971)

	Rupiah (10^6)			U.S.$ Equiv. (10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchange Portion	Local Currency Portion	Total
1. Equipment						
Plant	4,066	-	4,066	9,797	-	9,797
Utilities Plant & Facilities	2,749	-	2,749	6,625	-	6,625
Auxiliary & Off-Site	641	-	641	1,544	-	1,544
Sub-Total	7,456	-	7,456	17,966	-	17,966
2. Construction Costs						
Installation	925	535	1,460	2,230	1,289	3,519
Civil Works	931	1,111	2,042	2,244	2,678	4,922
Office Bldg. & Housing Cology	302	768	1,070	728	1,850	2,578
Sub-Total	2,158	2,414	4,572	5,202	5,817	11,019
3. Transportation	2,013	-	2,013	4,850	-	4,850
4. Plant Cost	11,627	2,414	14,041	28,018	5,817	33,835
5. Escalation & Contingency	7,932	1,647	9,579	19,114	3,969	23,083
6. Total Plant Costs	19,559	4,061	23,620	47,132	9,786	56,918
(1. Process Plant)	10,241	900	11,141	24,679	2,169	26,848
(2. Auxiliary & Off-Site)	8,810	1,870	10,680	21,228	4,506	25,734
(3. Buildings)	508	1,291	1,799	1,225	3,111	4,336
7. Royalty, Eng. Fee, Technical Exp.	4,195	-	4,195	10,108	-	10,108
8. Pre-Operating Expenses	-	1,150	1,150	-	2,770	2,770
9. Land	-	640	640	-	1,542	1,542
10. Interest During Construct on	1,943	-	1,943	4,682	-	4,682
11. Working Capital	-	9,268	9,268	-	22,332	22,332
12. Total Capital Requirement	25,697	15,119	40,816	61,922	36,430	98,352

Table IX-14 Estimated Capital Requirement (1977) Case D-1

Case D-1 p-TPA 70,000 t/y
DMT(W) 77,000 t/y

(Base Year 1971)

	Rupiah (10^6)			U.S. $ Equiv. (10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchange Portion	Local Currency Portion	Total
1. Equipment						
Plant	7,185	-	7,185	17,314	-	17,314
Utilities Plant & Facilities	3,141	-	3,141	7,569	-	7,569
Auxiliary & Off-Site	820	-	820	1,975	-	1,975
Sub-Total	11,146	-	11,146	26,858	-	26,858
2. Construction Costs						
Installation	1,461	800	2,261	3,520	1,928	5,448
Civil Works	1,252	1,467	2,719	3,017	3,536	6,553
Office Bldg. & Housing Colony	356	1,003	1,359	858	2,417	3,275
Sub-Total	3,069	3,270	6,339	7,395	7,881	15,276
3. Transportation	3,010	-	3,010	7,253	-	7,253
4. Plant Cost	17,225	3,270	20,495	41,506	7,881	49,387
5. Escalation & Contingency	11,750	2,231	13,981	28,314	5,375	33,689
6. Total Plant Costs	28,975	5,501	34,476	69,820	13,256	83,076
(1. Process Plant)	17,807	1,345	19,152	42,909	3,242	46,151
(2. Auxiliary & Off-Site)	10,569	2,468	13,037	25,467	5,947	31,414
(3. Buildings)	599	1,688	2,287	1,444	4,067	5,511
7. Royalty, Eng. Fee, Technical Exp.	7,440	-	7,440	17,928	-	17,928
8. Pre-Operating Expenses	-	1,397	1,397	-	3,365	3,365
9. Land	-	996	996	-	2,400	2,400
10. Interest During Construction	2,907	-	2,907	7,006	-	7,006
11. Working Capital	-	11,775	11,775	-	28,373	28,373
12. Total Capital Requirement	39,322	19,669	58,991	94,754	47,394	142,148

Table IX-15 Estimated Capital Requirement (1977) Case D-2

Case D-2 p-TPA 100,000 t/y
DMT(W) 110,000 t/y

(Base Year 1971)

	Rupiah (10^6)			U.S. $ Equiv. (10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchange Portion	Local Currency Portion	Total
1. Equipment						
Plant	8,510	-	8,510	20,506	-	20,506
Utilities Plant & Facilities	3,753	-	3,753	9,044	-	9,044
Auxiliary & Off-Site	959	-	959	2,311	-	2,311
Sub-Total	13,222	-	13,222	31,861	-	31,861
2. Construction Costs						
Installation	1,685	949	2,634	4,059	2,286	6,345
Civil Works	1,484	1,721	3,205	3,575	4,147	7,722
Office Bldg. & Housing Colony	437	1,084	1,521	1,053	2,611	3,664
Sub-Total	3,606	3,754	7,360	8,687	9,044	17,731
3. Transportation	3,570	-	3,570	8,603	-	8,603
4. Plant Cost	20,398	3,754	24,152	49,151	9,044	58,195
5. Escalation & Contingency	13,915	2,560	16,475	33,531	6,169	39,700
6. Total Plant Costs	34,313	6,314	40,627	82,682	15,213	97,895
(1. Process Plant)	21,014	1,595	22,609	50,635	3,844	54,479
(2. Auxiliary & Off-Site)	12,564	2,896	15,460	30,275	6,977	37,252
(3. Buildings)	735	1,823	2,558	1,772	4,392	6,164
7. Royalty, Eng. Fee, Technical Exp.	9,959	-	9,959	23,998	-	23,998
8. Pre-Operating Expenses	-	1,995	1,995	-	4,808	4,808
9. Land	-	1,176	1,176	-	2,833	2,833
10. Interest During Construction	3,528	-	3,528	8,500	-	8,500
11. Working Capital	-	16,271	16,271	-	39,208	39,208
12. Total Capital Requirement	47,800	25,756	73,556	115,180	62,062	177,242

The above estimate has been compiled firstly on the basis of an assumption that the construction would be undertaken in Indonesia in the year 1971, then the escalation and contingency factors were added thereto in order to obtain the total plant cost for the year 1976. Concerning the transportation, the freight from Japan to Indonesia including the marine insurance premium was calculated and the interest incurred during the construction period, royalty, engineering fee, land cost and working capital were added in order to obtain the total investment cost for the year 1977. Concerning the computation method for each of these cost items, a detailed explanation has already been made in the chapter concerning the Survey Method.

A brief explanation will be made here concerning the cases summarized in Table IX-7 through IX-15. The case A, pertains to the production of a large amount of c-TPA, approximately one-half of which will be retained as p-TPA and the balance is converted into DMT. Case B presupposes the employment of the Amoco p-TPA process alone, while case C employs only the Witten Process. Case D represents an instance in which the construction of the Amoco p-TPA Process and the Witten DMT Process plants are both carried out within the same plant premises under the common-use utility facilities arrangement.

5-3 Number of Personnel

Table IX-16 shows the direct and indirect number of people necessary for carrying out the operation. The total required number of workers is 346 for the p-TPA 100,000 t/y production, 346 for a Witten Process DMT 110,000 t/y production and 393 for the production of total 105,000 t/y of p-TPA and DMT via c-TPA. In the event of production scale alterations the number of workers will be slightly changed, however, the total required number of

Table IX-16 Required Number of Workers for p-TPA Plant (Direct, Indirect and Total)

(Person)

	p-TPA 100,000 t/y	Witten DMT 110,000 t/y	p-TPA 50,000 t/y DMT 55,000 "
Works Manager	1	1	1
Manager	2	2	2
Unit Superintendent	7	7	7
Section Superintendent	22	20	23
Foreman	52	47	59
Operator	161	153	190
Labourer	101	116	111
Total	346	346	393

personnel will not be greatly changed.

5-4 Operation Training

For the operation of a p-TPA or a DMT plant of 100,000 t/y scale, one unit superintendent, 12 section superintendents and 16 foreman, totalling 29 persons consisting of expatriate experts will be necessary to be allocated for approximately one year.

5-5 Plant Layout

Figure IX-5 shows the plant layout taken as an example of a p-TPA plant of 100,000 t/y capacity. The total area required for the premises excluding the housing colony, etc. is 92,500 m^2.

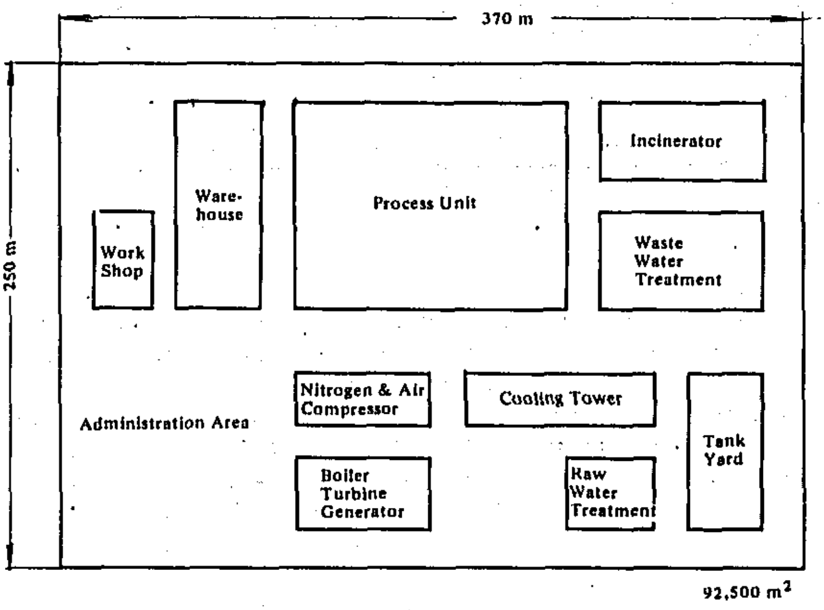


Figure IX-5 Plant Layout of 100,000 t/y p-TPA Plant

6. Economic Evaluation

6-1 Production and Operational Rate

It is assumed that the commencement of test operation will be undertaken in July 1977 and the production amount for the first year including a three-months test operation period shall be deemed as being 1/4 of the full production capacity. It is further assumed that a full capacity operation will be undertaken from the second year onward.

6-2 The Amount of Shipments to the Domestic Market and the Export Market

The extent of domestic demand has already been shown. Table IX-17 shows the total demand in terms of the demand for p-TPA or for DMT. Also, it is assumed here that the portion which utilizes p-TPA and the portion which utilizes DMT comprised in the total polyester demand are 1/2 each respectively in order to calculate the extent of the demand for p-TPA and DMT by taking into consideration the difference in the unit consumption for p-TPA and DMT at the time of polyester production. The results are shown in the same Table IX-17. Further, the total amount of the p-TPA and DMT demand is stipulated.

Table IX-17 Demand for p-TPA and DMT in Indonesia

(10^3 t/y)

		1977	1978	1979	1980	1981	1982	1983	1984	1985	1986
I	Total Demand (as p-TPA)	80	110	140	170	182	194	206	218	230	242
II	Total Demand (as DMT)	92	126	160	195	209	222	236	250	264	277
III	p-TPA Demand	40	55	70	85	91	97	103	109	115	121
IV	DMT Demand	46	63	80	98	105	111	118	125	132	139
V	Total Demand (as p-TPA, DMT)	86	118	150	183	196	208	221	234	247	260

Notes: III = I x 1/2
IV = II x 1/2
V = III + IV

This demand amount will be utilized as the basis for the case of simultaneous production of p-TPA and DMT. On the basis of the relationship between the domestic demand and production amount, the shipment extent to be made to the domestic market has been obtained and, the export amount has been calculated by subtracting the domestic demand from production amount. The obtained results are summarized in Table IX-18. The following paragraphs explain the case in which the simultaneous production of 100,000 t/y of p-TPA and 110,000 t/y of DMT is undertaken.

Table IX-18 Production, Domestic Shipment and Export Shipment of p-TPA and DMT

(10^3 t/y)

Case	Capacity		1977	1978	1979	1980	1981	1982	1983	1984	1985	1986
A-1	p-TPA 50	Production	26.25	105	105	105	105	105	105	105	105	105
	DMT 55	Domestic Consumption	26.25	105	105	105	105	105	105	105	105	105
		Export	0	0	0	0	0	0	0	0	0	0
A-2	p-TPA 70	Production	36.75	147	147	147	147	147	147	147	147	147
	DMT 77	Domestic Consumption	36.75	118	147	147	147	147	147	147	147	147
D-1	p-TPA 70 W-DMT 77	Export	0	29	0	0	0	0	0	0	0	0
A-3	p-TPA 100	Production	52.5	210	210	210	210	210	210	210	210	210
	DMT 110	Domestic Consumption	43	118	150	183	196	208	210	210	210	210
D-2	p-TPA 100 W-DMT 110	Export	9.5	92	60	27	14	2	0	0	0	0
		Production	17.5	70	70	70	70	70	70	70	70	70
B-1	p-TPA 70	Domestic Consumption	17.5	55	70	70	70	70	70	70	70	70
		Export	0	15	0	0	0	0	0	0	0	0
		Production	25	100	100	100	100	100	100	100	100	100
B-2	p-TPA 100	Domestic Consumption	20	55	70	85	91	97	100	100	100	100
		Export	5	45	30	15	9	3	0	0	0	0
		Production	19.25	77	77	77	77	77	77	77	77	77
C-1	W-DMT 77	Domestic Consumption	19.25	63	77	77	77	77	77	77	77	77
		Export	0	14	0	0	0	0	0	0	0	0
		Production	27.5	110	110	110	110	110	110	110	110	110
C-2	W-DMT 110	Domestic Consumption	23	63	80	98	105	110	110	110	110	110
		Export	4.5	47	30	12	5	0	0	0	0	0

The domestic demand in 1977 is assumed to be 86,000 t/y, however, because of the fact that the commencement of the production operation will be undertaken in July 1977, it would be possible to supply the products to the extent of 43,000 t/y which corresponds to the extent of demand for the period from July 1977 onward. Therefore, the domestic shipment will amount to 43,000 t/y and the remaining 9,500 t/y will be allocated for exportation. As has been explained in the above, the priority is given to the fulfillment of domestic demand. Concerning the demand from the year 1982 onward, the increment amount which would be attained from 1980 to 1981 has directly been applied to the relative calculations.

6-3 Production Cost Calculation

The relative calculation in this respect for obtaining the production cost have been undertaken by utilizing the standard unit consumption shown in Table IX-3 through IX-6, the investment amounts in Table IX-7 through IX-15 and the price factors summarized in the clause pertaining to Price Forecast. In order to carry out comparison studies between and among the various production processes, the relative comparisons will be made by taking the cases of producing the identical extent of production. In other words, three separate cases were assumed, i.e., in the case of producing 50,000 t/y of c-TPA and 55,000 t/y of DMT, totalling 105,000 t/y of the products via c-TPA, the production of 100,000 t/y of p-TPA alone and the production of 110,000 t/y of Witten Process DMT alone.

Table IX-19 through IX-21 show the production costs thus obtained. Figures IX-6 through IX-8 show the extent of variation caused in the production cost in the event that a ±20% fluctuation took place in each of the variable cost and in the total plant cost.

The lowest level of production cost will be achieved by the case of producing 100,000 t/y of p-TPA, while a slightly higher extent of production cost will be necessary for the case of simultaneous production of p-TPA and DMT via c-TPA. In the case of the Witten Process DMT, the production cost will become considerably high. The higher portion of the production cost in Witten Process is mostly due to the increase in utility cost (fuel cost). In the cost calculation shown in Table IX-19 through IX-21, it was assumed that heavy oil will be employed as the fuel. However, if a low cost natural gas can be utilized, the cost excess will be reduced. A further detailed explanation concerning this point will be made in the clause pertaining to the evaluation on the basis of the DCF method.

The effect of the variable cost variation upon the production cost is considerably high, however, that of the total plant cost would be comparatively low. In the case of a caprolactam plant, however, the effect on the total plant cost in the present case was considerably high. The extent of effect upon the production cost exerted by the variation in the variable cost and the total plant cost will be approximately on the same level for the three processes. In other words, Figures IX-6 through IX-8 can be approximately falling along one another by means of moving them in parallel.

6-4 Evaluation by the DCF Method

Table IX-22 and Figure IX-9 show the summary of the IRR concerning the 9 cases described in the foregoing. The highest extent of IRR is shown by the case A in which the production of p-TPA and DMT is undertaken via c-TPA, followed

Table IX-19 Production Cost of p-TPA and DMT (147,000 t/y)

Table IX-19-1 Estimated Production Cost of P-TPA'DMT(Amoco)

	Unit Consumption (kg/kg)	Price (US¢/kg)	Unit Cost (US¢/kg)
Variable Costs			
Raw Materials			
p-Xylene	0.632	42.6	26.92
Acetic Acid	0.0824	42.8	3.52
Methanol	0.194	20.8	4.03
Catalysts and Chemicals			0.43
Total Raw Material Costs			34.92
Utilities			
Fuel	0.753	8.3	6.24
Total Utility Costs			6.24
Total Variable Costs			41.17
Wages			0.49
Depreciation			
Battery Limits			2.79
Off-Site			1.20
Building			0.10
Royalty, Eng. Fee			2.64
Pre-Operating Expenses			0.41
Int. During Construction			0.86
Total Depreciation			8.02
Repairs and Insurance			1.96
Taxes			0
Plant Overhead Costs			0.49
Total Fixed Costs			10.98
Running Royalty			0
Factory Costs			52.15
Selling Expenses			1.40
General Administrative Exp.			1.56
Total Product. Cost Before Int.			55.12
Interest			
Int. on Total Inv. Costs			1.67
Int. on Working Capital			2.08
Total Interest			3.75
Total Product. Cost Including Int.			58.88

Working Capital (10^3 US$)	25,557
Total Investment Cost*(10^3 US$)	128,561

* Including Land Price & Working Capital

Table IX-19-2 Production Cost

(10³ US$/Y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production (10^3 ton)	36.75	147	147	147	147	147	147	147	147	147
Variable Costs										
Raw Materials										
p-Xylene	9,894	42,347	45,311	48,483	51,877	55,508	59,394	63,552	68,000	72,760
Acetic Acid	1,296	5,547	5,935	6,350	6,795	7,271	7,780	8,324	8,907	9,531
Methanol	1,482	6,346	6,791	7,266	7,775	8,319	8,901	9,525	10,191	10,905
Catalysts and Chemicals	160	687	735	786	842	900	964	1,031	1,103	1,181
Total Raw Material Costs	12,833	54,929	58,774	62,888	67,290	72,000	77,040	82,433	88,203	94,378
Utility										
Fuel	2,296	9,830	10,518	11,254	12,042	12,885	13,787	14,752	15,785	16,890
Total Utility Costs	2,296	9,830	10,518	11,254	12,042	12,885	13,787	14,752	15,785	16,890
Total Variable Costs	15,130	64,759	69,292	74,143	79,333	84,886	90,828	97,186	103,989	111,268
Wages	734	785	840	899	962	1,029	1,101	1,178	1,261	1,349
Depreciation										
Battery Limits	4,111	4,111	4,111	4,111	4,111	4,111	4,111	4,111	4,111	4,111
Off-Site	1,764	1,764	1,764	1,764	1,764	1,764	1,764	1,764	1,764	1,764
Building	157	157	175	157	157	157	157	157	157	157
Royalty, Eng. Fee	3,883	3,883	3,883	3,883	3,883	0	0	0	0	0
Pre-Operating Expenses	605	605	605	605	605	0	0	0	0	0
Int. During Construction	1,268	1,268	1,268	1,268	1,268	0	0	0	0	0
Total Depreciation	11,790	11,790	11,790	11,790	11,790	6,033	6,033	6,033	6,033	6,033
Repairs and Insurance	2,892	2,892	2,892	2,892	2,892	2,892	2,892	2,892	2,892	2,892
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	734	785	840	899	962	1,029	1,101	1,178	1,261	1,349
Total Fixed Costs	16,150	16,253	16,363	16,481	16,606	10,984	11,128	11,283	11,448	11,624
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	31,281	81,012	85,656	90,624	95,940	95,871	101,957	108,469	115,437	122,893
Selling Expenses	514	2,202	2,356	2,521	2,697	2,886	3,088	3,305	3,536	3,783
General Administrative Exp.	938	2,430	2,569	2,718	2,878	2,876	3,058	3,254	3,463	3,686
Total Product. Cost Before Int.	32,734	85,645	90,581	95,864	101,515	101,633	108,104	115,028	122,437	130,364
Interest										
Int. on Total Inv. Costs	4,915	4,915	4,915	4,915	4,213	3,511	2,809	2,106	1,404	702
Int. on Working Capital	3,066	2,044	1,022	0	0	0	0	0	0	0
Total Interest	7,982	6,960	5,938	4,915	4,213	3,511	2,809	2,106	1,404	702
Total Product. Cost Including Int.	40,717	92,605	96,520	100,779	105,729	105,145	110,913	117,135	123,841	131,066
(Unit Production Cost) (US¢/kg)	110.79	62.99	65.65	68.55	71.92	71.52	75.45	79.68	84.24	89.16

Table IX-19-3 Unit Production Cost

(US¢/kg)

Plant Life (Years)	1	2	3	4	5	6	7	8	9	10
Year (Year)	1977	1978	1979	1980	1981	1982	1983	1984	1985	1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production (10^3 ton)	36.75	147	147	147	147	147	147	147	147	147
Variable Costs										
Raw Materials										
p-Xylene	26.92	28.80	30.82	32.98	35.29	37.76	40.40	43.23	46.25	49.49
Acetic Acid	3.52	3.77	4.03	4.32	4.62	4.94	5.29	5.66	6.05	6.48
Methanol	4.03	4.31	4.61	4.94	5.28	5.65	6.05	6.47	6.93	7.41
Catalysts and Chemicals	0.43	0.46	0.50	0.53	0.57	0.61	0.65	0.70	0.75	0.80
Total Raw Material Costs	34.92	37.36	39.98	42.78	45.77	48.98	52.40	56.07	60.00	64.20
Utility										
Fuel	6.24	6.68	7.15	7.65	8.19	8.76	9.37	10.03	10.73	11.49
Total Utility Costs	6.24	6.68	7.15	7.65	8.19	8.76	9.37	10.03	10.73	11.49
Total Variable Costs	41.17	44.05	47.13	50.43	53.96	57.74	61.78	66.11	70.74	75.69
Wages	1.99	0.53	0.57	0.61	0.65	0.70	0.74	0.80	0.85	0.91
Depreciation										
Battery Limits	11.18	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79	2.79
Off-Site	4.80	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20
Building	0.42	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10
Royalty, Eng. Fee	10.56	2.64	2.64	2.64	2.64	0	0	0	0	0
Pre-Operating Expenses	1.64	0.41	0.41	0.41	0.41	0	0	0	0	0
Int. During Construction	3.45	0.86	0.86	0.86	0.86	0	0	0	0	0
Total Depreciation	32.08	8.02	8.02	8.02	8.02	4.10	4.10	4.10	4.10	4.10
Repairs and Insurance	7.87	1.96	1.96	1.96	1.96	1.96	1.96	1.96	1.96	1.96
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	1.99	0.53	0.57	0.61	0.65	0.70	0.74	0.80	0.85	0.91
Total Fixed Costs	43.94	11.05	11.13	11.21	11.29	7.47	7.57	7.67	7.78	7.90
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	85.11	55.11	58.26	61.64	65.26	65.21	69.35	73.78	78.52	83.60
Selling Expenses	1.40	1.49	1.60	1.71	1.83	1.96	2.10	2.24	2.40	2.57
General Administrative Exp.	2.55	1.65	1.74	1.84	1.95	1.95	2.08	2.21	2.35	2.50
Total Product. Cost Before Int.	89.07	58.26	61.62	65.21	69.05	69.13	73.54	78.25	83.29	88.68
Interest										
Int. on Total Inv. Costs	13.37	3.34	3.34	3.34	2.86	2.38	1.91	1.43	0.95	0.47
Int. on Working Capital	8.34	1.39	0.69	0	0	0	0	0	0	0
Total Interest	21.72	4.73	4.03	3.34	2.86	2.38	1.91	1.43	0.95	0.47
Total Product. Cost Including Int.	110.79	62.99	65.65	68.55	71.92	71.52	75.45	79.68	84.24	89.16

Table IX-19-4 Sales Revenue

(10³ US$/y)

Plant Life (Years)	1	2	3	4	5	6	7	8	9	10
Year (Year)	1977	1978	1979	1980	1981	1982	1983	1984	1985	1986
Domestic										
Sales Price (US¢/kg)	73	78	83	89	95	102	109	117	125	134
Sales Amount (ton)	36,750	118,000	147,000	147,000	147,000	147,000	147,000	147,000	147,000	147,000
Sales Value	26,901	92,422	123,195	131,819	141,046	150,920	161,484	172,788	184,883	197,825
Export										
Sales Price (US¢/kg)	52	55	59	63	68	73	78	83	89	95
Sales Amount (ton)	0	29,000	0	0	0	0	0	0	0	0
Sales Value	0	16,197	0	0	0	0	0	0	0	0
Total Sales Value	26,901	108,619	123,195	131,819	141,046	150,920	161,484	172,788	184,883	197,825

Table IX-19-5 Discounted Cash Flow Calculation

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Investment	128,561									
Land	1,908									
Inv. Excl. Land	126,653									
Total Sales Value	26,901	108,619	123,195	131,819	141,046	150,920	161,484	172,788	184,883	195,825
Total Product. Cost*	40,717	92,605	96,520	100,779	105,729	105,145	110,913	117,135	123,841	131,066
Income Before Tax	-13,816	16,014	26,675	31,040	35,317	45,775	50,571	55,653	61,042	64,759
Income Tax	0	0	0	0	0	20,598	22,756	25,043	27,468	29,141
Income After Tax	-13,816	16,014	26,675	31,040	35,317	25,176	27,814	30,609	33,573	35,617
Depreciation & Int.	19,772	18,750	17,728	16,706	16,003	9,544	8,842	8,140	7.437	6,735
Net Cash Flow (NCF)	5,956	34,764	44,403	47,746	51,320	34,720	36,656	38,749	41,010	42,352
Salvage Value (SV)	0	0	0	0	0	0	0	0	0	37,534
NCF + SV	5,956	34,764	44,403	47,746	51,320	34,720	36,656	38,749	41,010	79,887
Discount Rate	0.804355	0.646987	0.520407	0.418592	0.336697	0.270824	0.217839	0.175220	0.140939	0.113365
Present Value	4,790	22,491	23,107	19,986	17,279	9,403	7,985	6,789	5,779	9,056
(Total Present Value)										126,669
Internal Rate of Return (%)										24.3232

*Including Interest

Table IX-20 Production Cost of p-TPA (70,000 t/y)

Table IX-20-1 Estimated Production Cost of p-TPA

	Unit Consumption (kg/kg)	Price (US¢/kg)	Unit Cost (US¢/kg)
Variable Costs			
Raw Materials			
p-Xylene	0.69	42.6	29.39
Acetic Acid	0.096	42.8	4.10
Catalysts and Chemicals			0.61
Total Raw Material Costs			34.11
Utility			
Fuel	0.80	8.3	6.64
Total Utility Costs			6.64
Total Variable Costs			40.75
Wages			0.91
Depreciation			
Battery Limits			3.49
Off-Site			1.63
Building			0.16
Royalty, Eng. Fee			2.91
Pre-Operating Expenses			0.40
Int. During Construction			1.08
Total Depreciation			9.69
Repairs and Insurance			2.57
Taxes			0
Plant Overhead Costs			0.91
Total Fixed Costs			14.10
Running Royalty			0
Factory Costs			54.80
Selling Expenses			1.40
General Administrative Exp.			1.64
Total Product. Cost Before Int.			57.90
Interest			
Int. on Total Inv. Costs			2.11
Int. on Working Capital			2.19
Total Interest			4.31
Total Product. Cost Including Int.			62.21

Working Capital (10^3 US$)	12,800
Total Investment Cost*(10^3 US$)	74,459

* Including Land Price & Working Capital

Table IX-20-2 Production Cost

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production	17.5	70	70	70	70	70	70	70	70	70
Variable Costs										
Raw Materials										
p-Xylene	5,143	22,016	23,557	25,206	26,970	28,858	30,878	33,040	35,353	37,827
Acetic Acid	719	3,077	3,292	3,523	3,770	4,033	4,316	4,618	4,941	5,287
Catalysts and Chemicals	106	456	488	523	559	598	640	685	733	785
Total Raw Material Costs	5,969	25,550	27,339	29,252	31,300	33,491	35,835	38,344	41,028	43,900
Utility										
Fuel	1,162	4,973	5,321	5,694	6,092	6,519	6,975	7,463	7,986	8,545
Total Utility Costs	1,162	4,973	5,321	5,694	6,092	6,519	6,975	7,463	7,986	8,545
Total Variable Costs	7,131	30,523	32,660	34,946	37,393	40,010	42,811	45,808	49,014	52,445
Wages	643	688	736	787	842	901	964	1,032	1,104	1,182
Depreciation										
Battery Limits	2,449	2,449	2,449	2,449	2,449	2,449	2,449	2,449	2,449	2,449
Off-Site	1,141	1,141	1,141	1,141	1,141	1,141	1,141	1,141	1,141	1,141
Building	113	113	113	113	113	113	113	113	113	113
Royalty, Eng. Fee	2,039	2,039	2,039	2,039	2,039	0	0	0	0	0
Pre-Operating Expenses	285	285	285	285	285	0	0	0	0	0
Int. During Construction	759	759	759	759	759	0	0	0	0	0
Total Depreciation	6,787	6,787	6,787	6,787	6,787	3,703	3,703	3,703	3,703	3,703
Repairs and Insurance	1,800	1,800	1,800	1,800	1,800	1,800	1,800	1,800	1,800	1,800
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	643	688	736	787	842	901	964	1,032	1,104	1,182
Total Fixed Costs	9,873	9,963	10,060	10,163	10,273	7,307	7,434	7,569	7,713	7,868
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	17,005	40,487	42,720	45,109	47,666	47,318	50,245	53,377	56,728	60,314
Selling Expenses	245	1,048	1,122	1,200	1,284	1,374	1,470	1,573	1,683	1,801
General Administrative Exp.	510	1,214	1,281	1,353	1,429	1,419	1,507	1,601	1,701	1,809
Total Product. Cost Before Int.	17,760	42,750	45,124	47,663	50,381	50,112	53,223	56,552	60,114	63,925
Interest										
Int. on Total Inv. Costs	2,962	2,962	2,962	2,962	2,539	2,116	1,693	1,269	846	423
Int. on Working Capital	1,536	1,024	512	0	0	0	0	0	0	0
Total Interest	4,498	3,986	3,474	2,962	2,539	2,116	1,693	1,269	846	423
Total Product. Cost Including Int.	22,259	46,737	48,599	50,626	52,920	52,228	54,916	57,822	60,960	64,348
(Unit Production Cost) (US¢/kg)	127.19	66.76	69.42	72.32	75.60	74.61	78.45	82.60	87.08	91.92

Table IX-20-3 Unit Production Cost

(US¢/kg)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production (10^3 ton)	17.5	70	70	70	70	70	70	70	70	70
Variable Costs										
Raw Materials										
p-Xylene	29.39	31.45	33.65	36.00	38.52	41.22	44.11	47.20	50.50	54.03
Acetic Acid	4.10	4.39	4.70	5.03	5.38	5.76	6.16	6.59	7.05	7.55
Catalysts and Chemicals	0.61	0.65	0.69	0.74	0.79	0.85	0.91	0.97	1.04	1.12
Total Raw Material Costs	34.11	36.50	39.05	41.78	44.71	47.84	51.19	54.77	58.61	62.71
Utility										
Fuel	6.64	7.10	7.60	8.13	8.70	9.31	9.96	10.66	11.40	12.20
Total Utility Costs	6.64	7.10	7.60	8.13	8.70	9.31	9.96	10.66	11.40	12.20
Total Variable Costs	40.75	43.60	46.65	49.92	53.41	57.15	61.15	65.44	70.02	74.92
Wages	3.67	0.98	1.05	1.12	1.20	1.28	1.37	1.47	1.57	1.68
Depreciation										
Battery Limits	13.99	3.49	3.49	3.49	3.49	3.49	3.49	3.49	3.49	3.49
Off-Site	6.52	1.63	1.63	1.63	1.63	1.63	1.63	1.63	1.63	1.63
Building	0.64	0.16	0.16	0.16	0.16	0.16	0.16	0.16	0.16	0.16
Royalty, Eng. Fee	11.65	2.91	2.91	2.91	2.91	0	0	0	0	0
Pre-Operating Expenses	1.63	0.40	0.40	0.40	0.40	0	0	0	0	0
Int. During Construction	4.33	1.08	1.08	1.08	1.08	0	0	0	0	0
Total Depreciation	38.78	9.69	9.69	9.69	9.69	5.29	5.29	5.29	5.29	5.29
Repairs and Insurance	10.28	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	3.67	0.98	1.05	1.12	1.20	1.28	1.37	1.47	1.57	1.68
Total Fixed Costs	56.42	14.23	14.37	14.51	14.67	10.43	10.62	10.81	11.01	11.24
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	97.17	57.83	61.02	64.44	68.09	67.59	71.77	76.25	81.04	86.16
Selling Expenses	1.40	1.49	1.60	1.71	1.83	1.96	2.10	2.24	2.40	2.57
General Administrative Exp.	2.91	1.73	1.83	1.93	2.04	2.02	2.15	2.28	2.43	2.58
Total Product. Cost Before Int.	101.48	61.07	64.46	68.09	71.97	71.58	76.03	80.76	85.87	91.32
Interest										
Int. on Total Inv. Costs	16.93	4.23	4.23	4.23	3.62	3.02	2.41	1.81	1.20	0.60
Int. on Working Capital	8.77	1.46	0.73	0	0	0	0	0	0	0
Total Interest	25.70	5.69	4.96	4.23	3.62	3.02	2.41	1.81	1.20	0.60
Total Product. Cost Including Int.	127.19	66.76	69.42	72.32	75.60	74.61	78.45	82.60	87.08	91.92

Table IX-20-4 Sales Revenue

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Domestic										
Sales Price (US¢/kg)	73	78	83	89	95	102	109	117	125	134
Sales Amount (ton)	17,500	55,000	70,000	70,000	70,000	70,000	70,000	70,000	70,000	70,000
Sales Value	12,810	43,078	58,664	62,771	67,165	71,866	76,897	82,280	88,039	94,202
Export										
Sales Price (US¢/kg)	52	55	59	63	68	73	78	83	89	95
Sales Amount (ton)	0	15,000	0	0	0	0	0	0	0	0
Sales Value	0	8,378	0	0	0	0	0	0	0	0
Total Sales Value	12,810	51,456	58,664	62,771	67,165	71,866	76,897	82,280	88,039	94,202

Table IX-20-5 Discounted Cash Flow Calculation

(10³ US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Investment	74,459									
Land	1,233									
Inv. Excl. Land	73,226									
Total Sales Value	12,810	51,456	58,664	62,771	67,165	71,866	76,897	82,280	88,039	94,202
Total Product. Cost*	22,259	46,737	48,599	50,626	52,920	52,228	54,916	57,822	60,960	64,348
Income Before Tax	-9,449	4,719	10,065	12,145	14,245	19,638	21,981	24,458	27,079	29,854
Income Tax	0	0	0	0	0	8,837	9,891	11,006	12,185	13,434
Income After Tax	-9,449	4,719	10,065	12,145	14,245	10,800	12,089	13,451	14,893	16,419
Depreciation & Int.	11,286	10,774	10,262	9,750	9,327	5,820	5,396	4,973	4,550	4,127
Net Cash Flow (NCF)	1,837	15,493	20,327	21,895	23,572	16,620	17,485	18,424	19,443	20,546
Salvage Value (SV)	0	0	0	0	0	0	0	0	0	20,770
NCF + SV	1,837	15,493	20,327	21,895	23,572	16,620	17,485	18,424	19,443	41,316
Discount Rate	0.842473	0.709762	0.597956	0.503762	0.424406	0.357551	0.301227	0.253776	0.213800	0.180121
Present Value	1,547	10,996	12,154	11,029	10,004	5,942	5,267	4,675	4,157	7,442
(Total Present Value)										73,217
Internal Rate of Return (%)										18.698

* Including Interest

Table IX-21 Production Cost of Witten Process DMT (77,000 t/y)

Table IX-21-1 Estimated Production Cost of DMT(Witten)

	Unit Consumption (kg/kg)	Price (US¢/kg)	Unit Cost (US¢/kg)
Variable Costs			
Raw Materials			
p-Xylene	0.66	42.6	28.11
Methanol	0.41	20.8	8.52
Catalysts and Chemicals			0.28
Total Raw Material Costs			36.92
Utility			
Fuel	1.62	8.3	13.44
Total Utility Costs			13.44
Total Variable Costs			50.37
Wages			0.79
Depreciation			
Battery Limits			2.99
Off-Site			1.91
Building			0.16
Royalty; Eng. Fee			2.00
Pre-Operating Expenses			0.50
Int. During Construction			1.02
Total Depreciation			8.61
Repairs and Insurance			2.54
Taxes			0
Plant Overhead Costs			0.79
Total Fixed Costs			12.75
Running Royalty			0
Factory Costs			63.12
Selling Expenses			1.40
General Administrative Exp.			1.89
Total Product. Cost Before Int.			66.41
Interest			
Int. on Total Inv. Costs			1.98
Int. on Working Capital			2.52
Total Interest			4.50
Total Product. Cost Including Int.			70.92

Working Capital (10^3 US$)	16,202
Total Investment Cost*(10^3 US$)	80,200

* Including Land Price & Working Capital

Table IX-21-2 Production Cost

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production (10^3 ton)	19.25	77	77	77	77	77	77	77	77	77
Variable Costs										
Raw Materials										
p-Xylene	5,412	23,164	24,786	26,521	28,377	30,364	32,489	34,764	37,197	39,801
Methanol	1,641	7,026	7,518	8,044	8,607	9,209	9,854	10,544	11,282	12,072
Catalysts and Chemicals	53	230	246	264	282	302	323	346	370	396
Total Raw Material Costs	7,107	30,421	32,551	34,829	37,267	39,876	42,667	45,654	48,850	52,270
Utility										
Fuel	2,588	11,078	11,853	12,683	13,571	14,521	15,537	16,625	17,789	19,034
Total Utility Costs	2,588	11,078	11,853	12,683	13,571	14,521	15,537	16,625	17,789	19,034
Total Variable Costs	9,696	41,499	44,404	47,513	50,839	54,397	58,205	62,280	66,639	71,304
Wages	613	655	701	750	803	859	919	984	1,053	1,126
Depreciation										
Battery Limits	2,307	2,307	2,307	2,307	2,307	2,307	2,307	2,307	2,307	2,307
Off-Site	1,472	1,472	1,472	1,472	1,472	1,472	1,472	1,472	1,472	1,472
Building	129	129	129	129	129	129	129	129	129	129
Royalty, Eng. Fee	1,546	1,546	1,546	1,546	1,546	0	0	0	0	0
Pre-Operating Expenses	387	387	387	387	387	0	0	0	0	0
Int. During Construction	788	788	788	788	788	0	0	0	0	0
Total Depreciation	6,633	6,633	6,633	6,633	6,633	3,910	3,910	3,910	3,910	3,910
Repairs and Insurance	1,962	1,962	1,962	1,962	1,962	1,962	1,962	1,962	1,962	1,962
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	613	655	701	750	803	859	919	984	1,053	1,126
Total Fixed Costs	9,821	9,907	9,999	10,097	10,202	7,592	7,713	7,841	7,979	8,127
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	19,518	51,407	54,404	57,610	61,042	61,990	65,918	70,121	74,619	79,431
Selling Expenses	269	1,153	1,234	1,320	1,413	1,512	1,617	1,731	1,852	1,982
General Administrative Exp.	585	1,542	1,632	1,728	1,831	1,859	1,977	2,103	2,238	2,382
Total Product. Cost Before Int.	20,373	54,103	57,270	60,659	64,286	65,362	69,514	73,956	78,710	83,796
Interest										
Int. on Total Inv. Costs	3,051	3,051	3,051	3,051	2,615	2,179	1,743	1,307	811	435
Int. on Working Capital	1,944	1,296	648	0	0	0	0	0	0	0
Total Interest	4,995	4,347	3,699	3,051	2,615	2,179	1,743	1,307	811	435
Total Product. Cost Including Int.	25,368	58,450	60,970	63,711	66,901	67,541	71,257	75,264	79,582	84,232
(Unit Production Cost) (US¢/kg)	131.78	75.90	79.18	82.74	86.88	87.71	92.54	97.74	103.35	109.39

Table IX-21-3 Unit Production Cost

(US¢/kg)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production	19.25	77	77	77	77	77	77	77	77	77
Variable Costs										
Raw Materials										
p-Xylene	28.11	30.08	32.19	34.44	36.85	39.43	42.19	45.14	48.30	51.69
Methanol	8.52	9.12	9.76	10.44	11.17	11.96	12.79	13.69	14.65	15.67
Catalysts and Chemicals	0.28	0.29	0.32	0.34	0.36	0.39	0.42	0.44	0.48	0.51
Total Raw Material Costs	36.92	39.50	42.27	45.23	48.39	51.78	55.41	59.29	63.44	67.88
Utility										
Fuel	13.44	14.38	15.39	16.47	17.62	18.85	20.17	21.59	23.10	24.71
Total Utility Costs	13.44	14.38	15.39	16.47	17.62	18.85	20.17	21.59	23.10	24.71
Total Variable Costs	50.37	53.89	57.66	61.70	66.02	70.64	75.59	80.88	86.54	92.60
Wages	3.18	0.85	0.91	0.97	1.04	1.11	1.19	1.27	1.36	1.46
Depreciation										
Battery Limits	11.98	2.99	2.99	2.99	2.99	2.99	2.99	2.99	2.99	2.99
Off-Site	7.65	1.91	1.91	1.91	1.91	1.91	1.91	1.91	1.91	1.91
Building	0.67	0.16	0.16	0.16	0.16	0.16	0.16	0.16	0.16	0.16
Royalty, Eng. Fee	8.03	2.00	2.00	2.00	2.00	0	0	0	0	0
Pre-Operating Expenses	2.01	0.50	0.50	0.50	0.50	0	0	0	0	0
Int. During Construction	4.09	1.02	1.02	1.02	1.02	0	0	0	0	0
Total Depreciation	34.45	8.61	8.61	8.61	8.61	5.07	5.07	5.07	5.07	5.07
Repairs and Insurance	10.19	2.54	2.54	2.54	2.54	2.54	2.54	2.54	2.54	2.54
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	3.18	0.85	0.91	0.97	1.04	1.11	1.19	1.27	1.36	1.46
Total Fixed Costs	51.02	12.86	12.98	13.11	13.25	9.86	10.01	10.18	10.36	10.55
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	101.39	66.76	70.65	74.81	79.27	80.50	85.60	91.06	96.90	103.15
Selling Expenses	1.40	1.49	1.60	1.71	1.83	1.96	2.10	2.24	2.40	2.57
General Administrative Exp.	3.04	2.00	2.11	2.24	2.37	2.41	2.56	2.73	2.90	3.09
Total Product. Cost Before Int.	105.83	70.26	74.37	78.77	83.48	84.88	90.27	96.04	102.22	108.82
Interest										
Int. on Total Inv. Costs	15.85	3.96	3.96	3.96	3.39	2.83	2.26	1.69	1.13	0.56
Int. on Working Capital	10.10	1.68	0.84	0	0	0	0	0	0	0
Total Interest	25.95	5.64	4.80	3.96	3.39	2.83	2.26	1.69	1.13	0.56
Total Product. Cost Including Int.	131.78	75.90	79.18	82.74	86.88	87.71	92.54	97.74	103.35	109.39

Table IX-21-4 Sales Revenue

(10^3 US$/y)

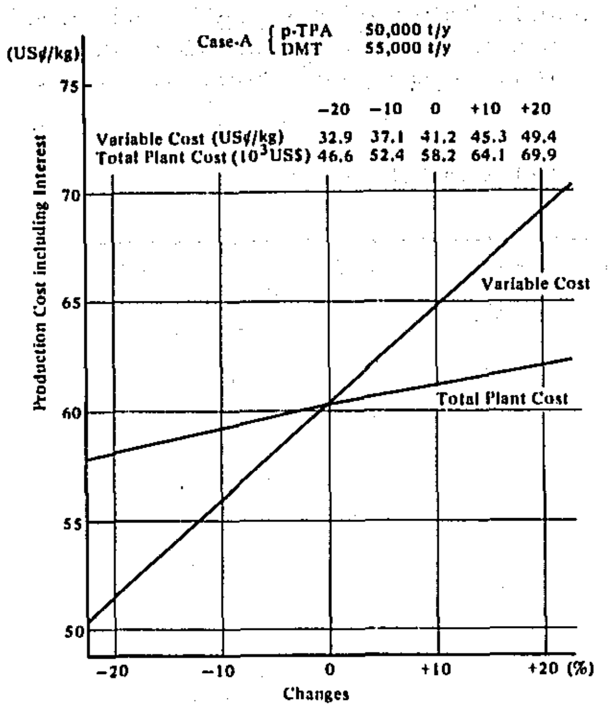
Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Domestic										
Sales Price (US¢/kg)	73	78	83	89	95	102	109	117	125	134
Sales Amount (ton)	19,250	63,000	77,000	77,000	77,000	77,000	77,000	77,000	77,000	77,000
Sales Value	14,091	49,344	64,531	69,048	73,881	79,053	84,587	90,508	96,843	103,622
Export										
Sales Price (US¢/kg)	52	55	59	63	68	73	78	83	89	95
Sales Amount (ton)	0	14,000	0	0	0	0	0	0	0	0
Sales Value	0	7,819	0	0	0	0	0	0	0	0
Total Sales Value	14,091	57,163	64,531	69,048	73,881	79,053	84,587	90,508	96,843	103,622

Table IX-21-5 Discounted Cash Flow Calculation

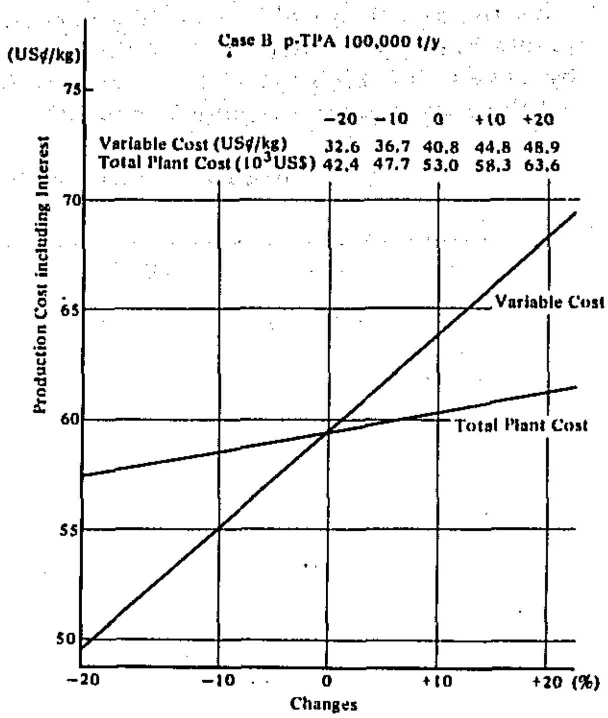
(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Investment	80,200									
Land	1,317									
Inv. Excl. Land	78,883									
Total Sales Value	14,091	57,163	64,531	69,048	73,881	79,053	84,587	90,508	96,843	103,622
Total Product. Cost*	25,368	58,450	60,970	63,711	66,901	67,541	71,257	75,264	79,582	84,232
Income Before Tax	-11,277	-1,287	3,561	5,337	6,980	11,512	13,330	15,244	17,261	19,390
Income Tax	0	0	0	0	0	5,180	5,998	6,859	7,767	8,725
Income After Tax	-11,277	-1,287	3,561	5,337	6,980	6,331	7,331	8,384	9,493	10,664
Depreciation & Int.	11,628	10,980	10,332	9,684	9,248	6,089	5,654	5,218	4,782	4,346
Net Cash Flow (NCF)	351	9,693	13,893	15,021	16,228	12,420	12,985	13,602	14,275	15,010
Salvage Value (SV)	0	0	0	0	0	0	0	0	0	26,165
NCF + SV	351	9,693	13,893	15,021	16,228	12,420	12,985	13,602	14,275	41,175
Discount Rate	0.905075	0.819161	0.741402	0.671024	0.607327	0.549676	0.497498	0.450273	0.407531	0.368846
Present Value	317	7,940	10,300	10,079	9,855	6,827	6,460	6,124	5,817	15,187
(Total Present Value)										78,910
Internal Rate of Return (%)										10.4881

* Including Interest



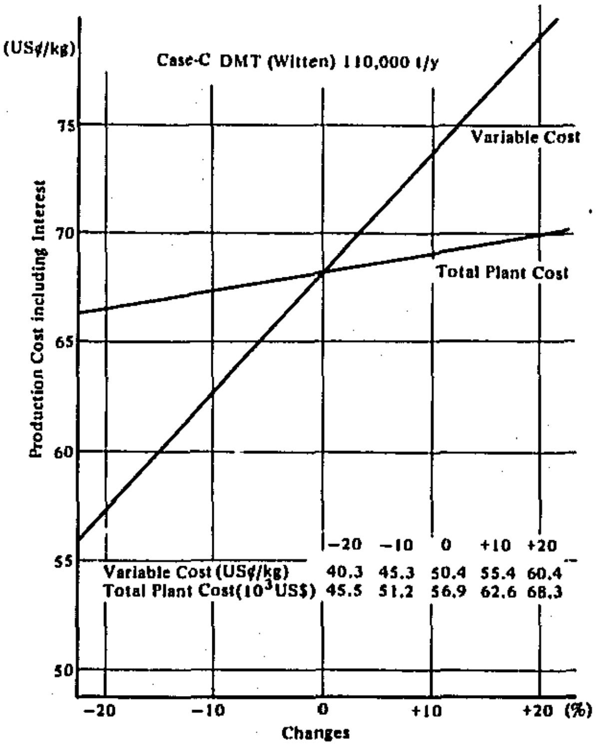
	−20	−10	0	+10	+20
Variable Cost (US¢/kg)	32.9	37.1	41.2	45.3	49.4
Total Plant Cost (10^3US$)	46.6	52.4	58.2	64.1	69.9



	−20	−10	0	+10	+20
Variable Cost (US¢/kg)	32.6	36.7	40.8	44.8	48.9
Total Plant Cost (10^3US$)	42.4	47.7	53.0	58.3	63.6

Figure IX-6 Sensitivity Analysis of Production Cost, Case A

Figure IX-7 Sensitivity Analysis of Production Cost, Case B



	−20	−10	0	+10	+20
Variable Cost (US¢/kg)	40.3	45.3	50.4	55.4	60.4
Total Plant Cost(10^3US$)	45.5	51.2	56.9	62.6	68.3

Figure IX-8 Sensitivity Analysis of Production Cost, Case C by case B in which the p-TPA production alone is undertaken. Then the third place goes to case D in which a p-TPA production and a Witten Process DMT production plants are to be constructed within the same premises. The lowest IRR is shown by case C in which the Witten Process DMT alone is produced. Except for case C, all the IRR figures are more than 15%.

Table IX-22 IRR of Various Cases of p-TPA and DMT Production

	p-TPA (10^3 t/y)	W-DMT (10^3 t/y)	c-TPA (10^3 t/y)		IRR (%)	Total Product. Cost Incl. Int. (US¢/kg)
			p-TPA	DMT		
A - 1			50	55	22.5	60.4
2			70	77	24.3	58.9
3			100	110	23.8	57.5
B - 1	70				18.7	62.2
2	100				19.8	59.5
C - 1		77			10.5	70.9
2		110			11.4	68.2
D - 1	70	77			16.8	65.2
2	100	110			17.5	62.9

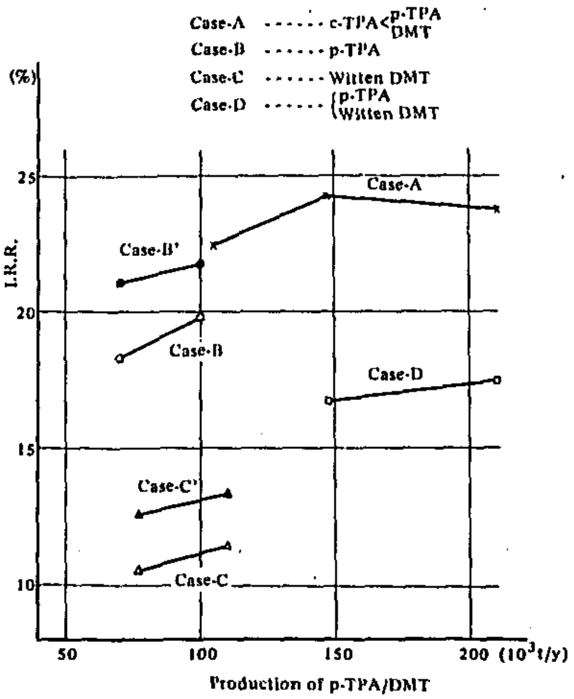


Figure IX-9 Production Amount of p-TPA, DMT and IRR

When p-TPA and DMT of 100,000 t/y is produced, case B shows a lower extent of production cost than case A; however, when compared on the basis of IRR, case A shows a higher level. The reason for such an inversion is that the export amount will be greater in the case of producing p-TPA alone due to the correlation with the domestic demand (refer to Table IX-18). In case A, approximately 1/2 each of p-TPA and DMT will be produced so that no low-cost export allocation will be made. Therefore, the IRR in this case will become higher. If it seems to be possible to confine the raw material demand for polyester fiber to p-TPA alone in Indonesia, the IRR of case B will become the greatest.

The case B' and case C' stipulated in Figure IX-9 represent the cases in which the calculation of IRR have been made by independently separating p-TPA and DMT produced in case D. In these cases, the IRR showed a figure approximately 2% greater when compared with cases B and C in which completely independent production has been taken, because of the fact that in case B' and case C', the utility facilities were designed to be commonly used. Figures IX-10 through IX-12 show results of a sensitivity analysis of IRR by assuming a production of 70,000 to 77,000 t/y each of p-TPA and DMT in cases A, B and C. Tables IX-19 through IX-21 show the results of calculation on IRR of these cases. Sales price is the factor which exerts the greatest influence on IRR, followed by the variable cost. The effect by the total plant cost is considerably low.

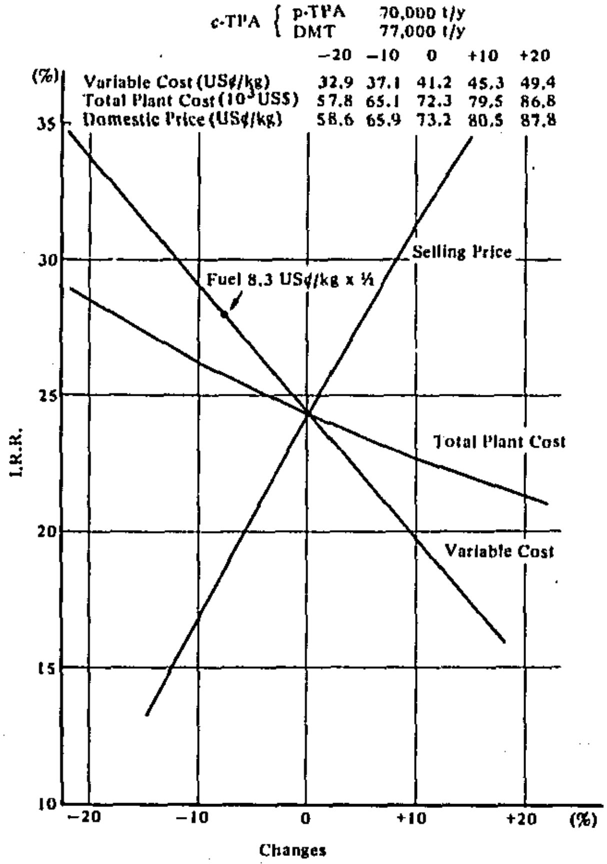


Figure IX-10 Sensitivity Analysis of IRR in p-TPA and DMT Production (Case A-2)

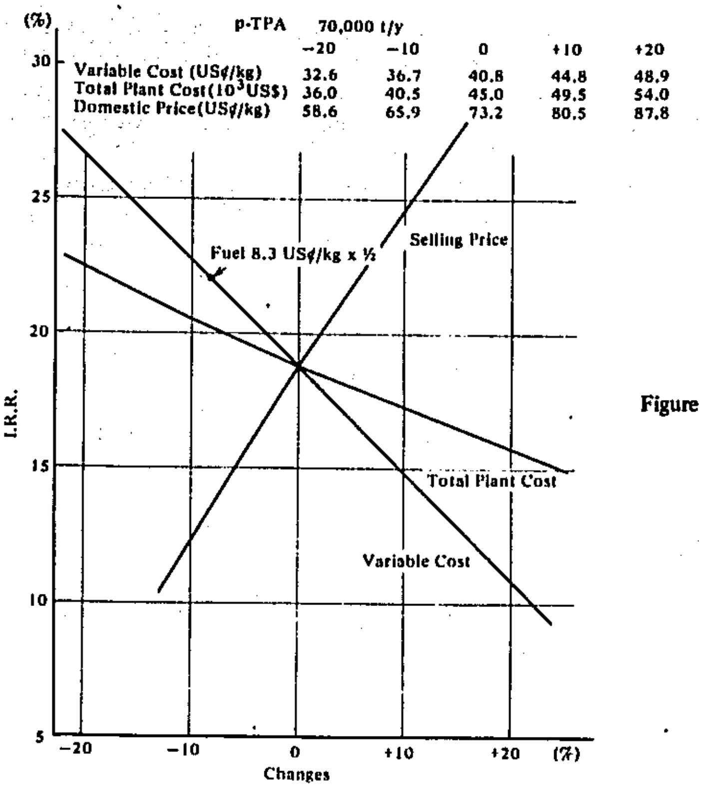


Figure IX-11 Sensitivity Analysis of IRR in p-TPA Production (Case B-1)

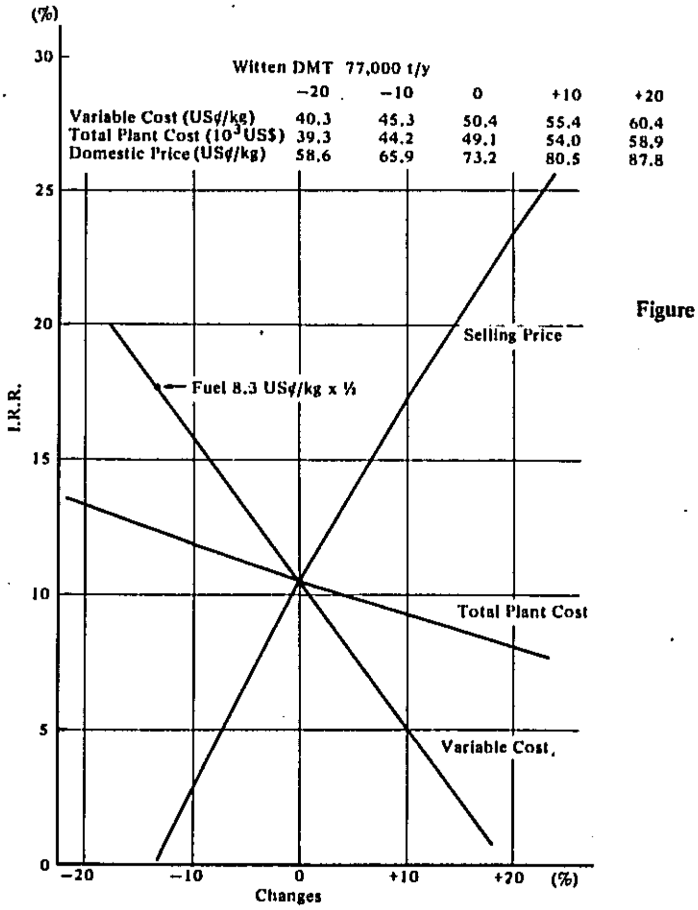


Figure IX-12 Sensitivity Analysis of IRR in DMT Production (Case C-1)

The IRR in the case of the Witten Process DMT is smaller when compared with the other processes. The major factor causing such a phenomenon is the fuel cost, as has been described in the clause pertaining to the production cost. Due to the fact that the conversion rate in the oxidation reaction is low in the case of the Witten Process, it is necessary to undertake a large amount of recycling and further, the crystallization process is required for refining, thereby causing a large extent of utility consumption. Also, because of the generation of the various impurities from the side-reactions, the consumption of p-xylene will become high. The theoretical unit consumption of p-xylene against DMT is 0.55, however, 0.66 is necessary for the Witten Process.

In the past cases where the crude oil price was low and, therefore, the fuel cost was low, the difference in the fuel cost unit consumption did not exert any significant effect on the cost. For instance, if it is assumed that the fuel price is 1/4 of US¢8.3/kg, the difference between the p-TPA and the Witten Process DMT will be drastically reduced from the present level of US¢6.8/kg to US¢1.7/kg. Therefore, if it is possible to employ low cost natural gas as the fuel, the difference in IRR between the p-TPA and Witten Process DMT will be reduced.

If the price of energy per unit in Japan for the liquified natural gas and crude oil seems to be on the same level, the per-energy-cost of natural gas in Indonesia could be assessed as being approximately 1/2 of that of crude oil. Therefore, if it is possible to utilize such natural gas, the difference in the fuel cost between p-TPA and Witten Process DMT will be US¢3.4/kg. In this case, the variable cost of the Witten Process DMT will be reduced by US¢6.7/kg, and in the case of 77,000 t/y production, the IRR will be drastically improved from 10.5% to 17.6%. Further, in this case, although the IRR of p-TPA will increase from 18.7% to 22.0%, the difference between the IRR of the p-TPA and the Witten Process DMT will nevertheless be considerably reduced.

The largest extent of IRR of all the cases shown is by case A. Figure IX-13 shows the effect on IRR exerted in the event where the export rate of case A is varied. In this figure, the term "Standard" signifies the case in which the priority is given to the domestic demand fulfillment which has been applied so far and, the items other than the "Standard" are the cases in which the constant allocation of 15% and 30% of the production is made to exportation.

As is evident from Figure IX-13, the IRR will be drastically reduced when the export rate is increased. However, except for the case of "Standard", the IRR becomes higher when the production is greater even if the export rate is maintained on the same level. In the case of the "Standard", the IRR is reduced between the range from 147,000 t/y to 210,000 t/y. This signifies that the IRR of the additional investment within this range is smaller than the 24.3% IRR which is the case of 147,000 t/y production.

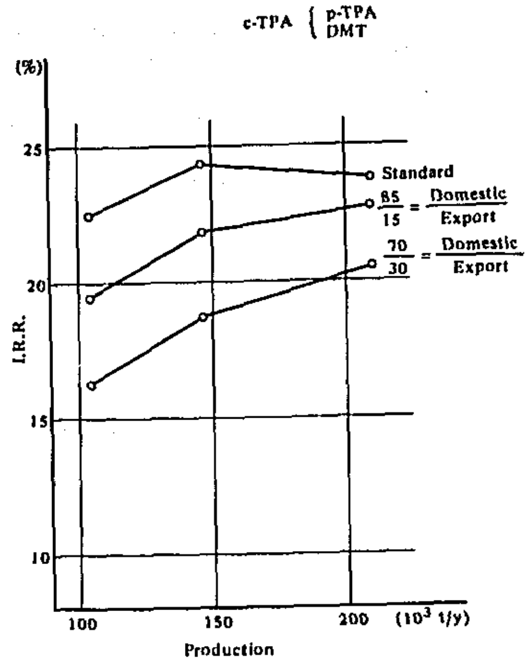


Figure IX-13 Export Rate of p-TPA, DMT and IRR (Case A)

6-5 Prices of p-TPA and DMT

Economic evaluation is undertaken on the assumption that the prices of p-TPA and DMT are identical with each other. However, as described in Chapter V 4-2-4, it is economically advantageous to employ p-TPA to produce polyester chip under the condition that the price of p-TPA is identical to the price of DMT. Therefore, it might be practical to evaluate the price of p-TPA higher than that of DMT. However, in trade, p-TPA and DMT have been sold at the same price on a weight base, therefore the prices of p-TPA and DMT are regarded as the same in the economic evaluation.

In the following paragraphs, the explanation is made on the price difference in p-TPA and DMT from the view point of production cost of polyester chip.

Pre-requisite Conditions:

(1) The quality of fibers derived from p-TPA is equivalent to in every way to that derived from DMT.

(2) Investment on polymerization facilities based on p-TPA is 5% less than that based on DMT. In the initial stage of polymerization, the required ethylene glycol per mole of p-TPA in p-TPA process is less than that of DMT process, therefore reduction is possible in the size of polymerization facilities in p-TPA process.

(3) Unit consumption is 0.86 for p-TPA and 1.0 for DMT and there are no differences in unit consumption of ethylene glycol and costs of catalysts & chemicals in polymerization processes based on the two monomers.

(4) Utility cost of polymerization based on DMT is 12% higher than that based on p-TPA.

(5) There are no differences in costs between the two processes other than the above.

On the above assumption, the factors which affect the correlation between prices of p-TPA, DMT and production cost of polyester chip are as follows.

(1) Prices of p-TPA and DMT

(2) Price of recovered methanol

The relation between these prices and production cost of polyester chip is shown in Figure IX-14 and IX-15.

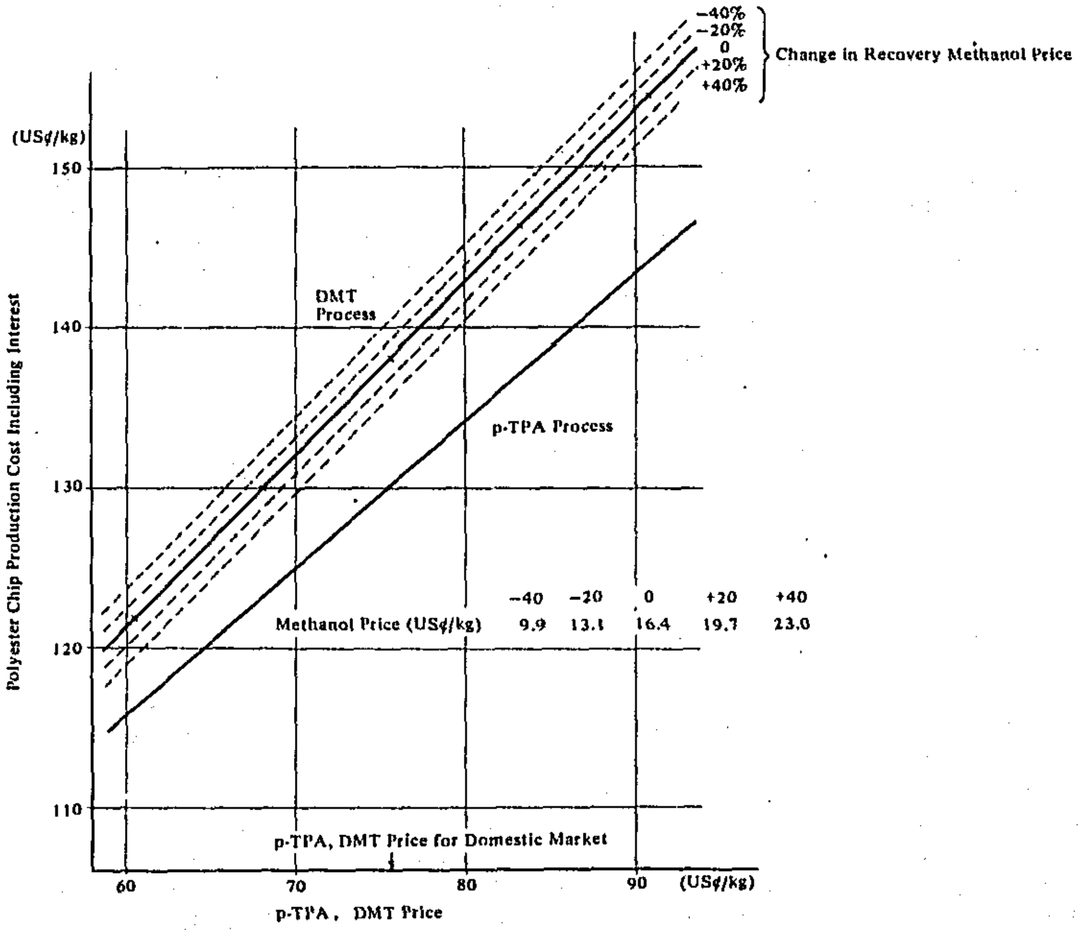


Figure IX-14 Price of Recovery Methanol, p-TPA, DMT and Production Cost of Polyester Chip

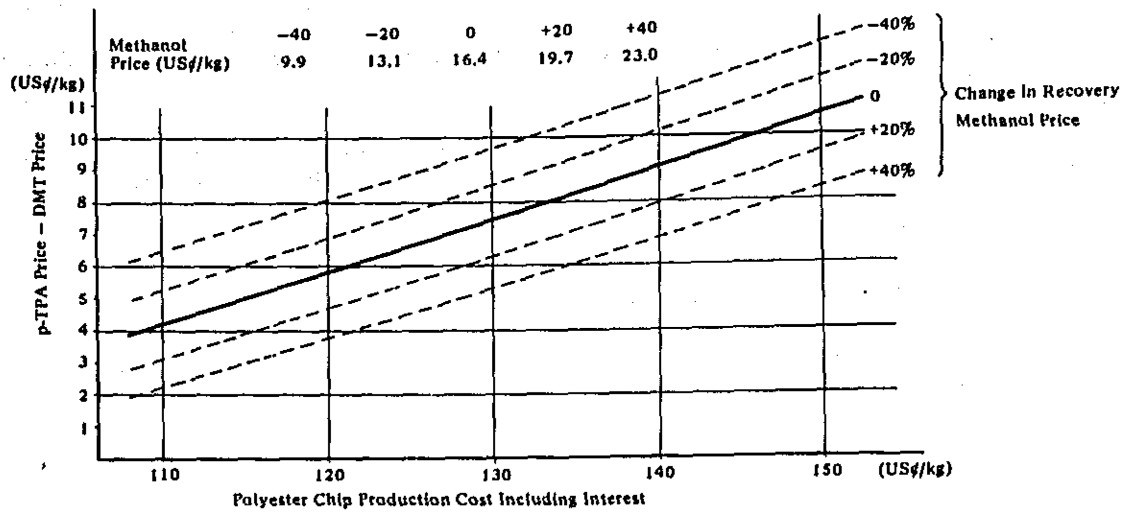


Figure IX-15 Production Cost of Polyester Chip, Recovery Methanol Price and Price Difference in p-TPA and DMT

As is evident from these Figures, production cost of polyester chip based on p-TPA is about US¢8/kg less than that based on DMT, when the prices of p-TPA and DMT are identical. Further, the price of recovered methanol must be twice of standard methanol price US¢16.4/kg to produce polyester chip from DMT at the same production cost as that from p-TPA. However, when the price of recovered methanol is high the price of methanol used to produce DMT is high, therefore it will be very difficult to produce DMT at the same production cost as that of p-TPA.

On an assumption that the prices of p-TPA and DMT are identical, as is evident from the above discussion, it is obviously advantageous for polyester manufacturers to select p-TPA.

7. Extent of Contribution to the Indonesian National Economy

The extent of contribution to the national economy of Indonesia by the implementation of this project was assessed in view of the saving in foreign currency by employing the shadow pricing method. The results are shown in Tables IX-23 and IX-24. The amount of saving attainable in foreign currency for a period of ten years from the implementation of the project was calculated on the basis of the shadow exchange rate of US$1 = Rp519. The results reveal, as shown in Table IX-23, that the amount will be US$573 million in the case of producing 70,000 t/y of p-TPA and 77,000 t/y of DMT. When this amount is discounted by 15%/y, the amount becomes US$258 million which implies a considerable contribution to the foreign currency saving of Indonesia.

Then, by employing the method described in the clause pertaining to the Survey Method, the economic evaluation in view of national economy was undertaken by utilizing the shadow exchange rate of US$1 = Rp519 and the labor cost on the basis of 40% of the actually prevailing wages. The results are shown in Table IX-24. IRR corresponds to 27% so that the implementation of this project will greatly contribute to the future development of Indonesia.

Also, one of the objectives of the Second Five-year Plan is the "The attainment of sufficient supply of food and clothing commensurate with the purchasing power of the people". Also in view of this objective, it can be stated that this project completely meets the above-mentioned objective.

Table IX-23 Foreign Currency Saving Attainable by 70,000 t/y p-TPA Production and 77,000 t/y DMT Production

(10^3 US$)

	A	B	C	D	E	F	Present Value
1977	23,979	9,126	5,182	0	1,563	8,108	7,050
1978	82,384	31,354	5,182	20,246	6,691	59,403	44,917
1979	109,815	41,793	5,182	0	7,159	55,681	36,511
1980	117,502	44,719	5,182	0	7,660	59,941	34,271
1981	125,728	47,849	19,917	0	8,197	49,765	24,741
1982	134,529	51,199	19,176	0	8,770	55,384	23,944
1983	143,946	54,783	18,436	0	9,384	61,343	23,061
1984	154,022	58,618	17,696	0	10,041	67,667	22,120
1985	164,804	62,721	16,956	0	10,744	74,383	21,144
1986	176,340	67,111	16,215	0	11,497	81,517	20,149
Total	1,233,049	469,273	129,124	20,246	81,706	573,192	258,008

Notes: A: Sales Value of Product
B: Export Value of Raw Materials
C: Repayment of Foreign Loan and Interest on Foreign Loan
D: Export Value of Product
E: Import Value of Sub-Raw Materials
F: $F = A - B - C + D - E$
Yearly Foreign Currency Saving
Present Value: $\frac{Fn}{(1+0.15)^n}$

Table IX-24 Evaluation of National Benefit by p-TPA and DMT Production (Amoco)

Table IX-24-1 Estimated Production Cost of p-TPA, DMT(Amoco) (147,000 t/y)

	Unit Consumption (kg/kg)	Price (US¢/kg)	Unit Cost (US¢/kg)
Variable Costs			
Raw Materials			
p-Xylene	0.632	44.75	28.28
Acetic Acid	0.0824	45	3.70
Methanol	0.194	21.875	4.24
Catalysts and Chemicals			0.54
Total Raw Material Costs			36.77
Utility			
Fuel	0.753	10.375	7.81
Total Utility Costs			7.81
Total Variable Costs			44.59
Wages			0.19
Depreciation			
Battery Limits			3.44
Off-Site			1.44
Building			0.11
Royalty, Eng. Fee			3.30
Pre-Operating Expenses			0.44
Int. During Construction			1.04
Total Depreciation			9.79
Repairs and Insurance			2.38
Taxes			0
Plant Overhead Costs			0.19
Total Fixed Costs			12.57
Running Royalty			0
Factory Costs			57.16
Selling Expenses			1.40
General Administrative Exp.			1.71
Total Product. Cost Before Int.			60.28
Interest			
Int. on Total Inv. Costs			2.03
Int. on Working Capital			2.28
Total Interest			4.31
Total Product. Cost Including Int.			64.60

Working Capital (10^3 US$)	28,013
Total Investment Cost *(10^3 US$)	152,683

* Including Land Price & Working Capital

Table IX-24-2 Production Cost

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production (10^3 ton)	36.75	147	147	147	147	147	147	147	147	147
Variable Costs										
Raw Materials										
p-Xylene	10,393	44,484	47,598	50,930	54,495	58,310	62,392	66,759	71,432	76,433
Acetic Acid	1,362	5,832	6,240	6,677	7,144	7,644	8,180	8,752	9,365	10,020
Methanol	1,559	6,674	7,142	7,642	8,177	8,749	9,362	10,017	10,718	11,468
Catalysts and Chemicals	200	859	919	983	1,052	1,126	1,204	1,289	1,379	1,476
Total Raw Material Costs	13,516	57,851	61,900	66,233	70,870	75,831	81,139	86,819	92,896	99,399
Utility										
Fuel	2,871	12,288	13,148	14,068	15,053	16,107	17,234	18,441	19,731	21,113
Total Utility Costs	2,871	12,288	13,148	14,068	15,053	16,107	17,234	18,441	19,731	21,113
Total Variable Costs	16,387	70,139	75,049	80,302	85,923	91,938	98,373	105,260	112,628	120,512
Wages	293	313	335	358	384	410	439	470	503	538
Depreciation										
Battery Limits	5,067	5,067	5,067	5,067	5,067	5,067	5,067	5,067	5,067	5,067
Off-Site	2,120	2,120	2,120	2,120	2,120	2,120	2,120	2,120	2,120	2,120
Building	168	168	168	168	168	168	168	168	168	168
Royalty, Eng. Fee	4,853	4,853	4,853	4,853	4,853	0	0	0	0	0
Pre-Operating Expenses	655	655	655	655	655	0	0	0	0	0
Int. During Construction	1,535	1,535	1,535	1,535	1,535	0	0	0	0	0
Total Depreciation	14,401	14,401	14,401	14,401	14,401	7,356	7,356	7,356	7,356	7,356
Repairs and Insurance	3,501	3,501	3,501	3,501	3,501	3,501	3,501	3,501	3,501	3,501
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	293	313	335	358	384	410	439	470	503	538
Total Fixed Costs	18,488	18,530	18,573	18,620	18,671	11,680	11,737	11,799	11,865	11,935
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	34,876	88,669	93,622	98,923	104,594	103,616	110,111	117,059	124,493	132,448
Selling Expenses	514	2,202	2,356	2,521	2,697	2,886	3,088	3,305	3,536	3,783
General Administrative Exp.	1,046	2,660	2,808	2,967	3,137	3,108	3,303	3,511	3,734	3,973
Total Product. Cost Before Int.	36,437	93,531	98,787	104,412	110,430	109,613	116,503	123,876	131,764	140,205
Interest										
Int. on Total Inv. Costs	5,970	5,970	5,970	5,970	5,117	4,264	3,411	2,558	1,705	852
Int. on Working Capital	3,361	2,241	1,120	0	0	0	0	0	0	0
Total Interest	9,331	8,211	7,090	5,970	5,117	4,264	3,411	2,558	1,705	852
Total Product. Cost Including Int.	45,769	101,742	105,878	110,382	115,547	113,878	119,915	126,434	133,470	141,058
(Unit Production Cost) (US¢/kg)	124.54	69.21	72.02	75.09	78.60	77.46	81.57	86.01	90.79	95.95

Table IX-24-3 Sales Revenue

(10³ US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Domestic										
Sales Price (US¢/kg)	79	85	91	97	104	111	119	127	77 6	146
Sales Amount (ton)	36,750	118,000	147,000	147,000	147,000	147,000	147,000	147,000	147,000	147,000
Sales Value	29,216	100,376	133,798	143,164	153,186	163,909	175,382	187,659	200,795	214,851
Export										
Sales Price (US¢/kg)	65	69	74	79	85	91	97	104	112	119
Sales Amount (ton)	0	29,000	0	0	0	0	0	0	0	0
Sales Value	0	20,247	0	0	0	0	0	0	0	0
Total Sales Value	29,216	120,623	133,798	143,164	153,186	163,909	175,382	187,659	200,795	214,851

Table IX-24-4 Discounted Cash Flow Calculation

(10³ US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Investment	152,683									
Land	1,908									
Inv. Excl. Land	150,775									
Total Sales Value	29,216	120,623	133,798	143,164	153,186	163,909	175,382	187,659	200,795	214,851
Total Product. Cost*	45,769	101,742	105,878	110,382	115,547	113,878	119,915	126,434	133,470	141,058
Income Before Tax	-16,553	18,881	27,920	32,782	37,639	50,031	55,467	61,225	67,325	73,793
Income Tax	0	0	0	0	0	0	0	0	0	0
Income After Tax	-16,553	18,881	27,920	32,782	37,639	50,031	55,467	61,225	67,325	73,793
Depreciation & Int.	23,733	22,612	21,492	20,371	19,518	11,621	10,768	9,915	9,062	8,209
Net Cash Flow (NCF)	7,180	41,493	49,412	53,153	57,157	61,652	66,235	71,140	76,387	82,002
Salvage Value (SV)	0	0	0	0	0	0	0	0	0	41,984
NCF + SV	7,180	41,493	49,412	53,153	57,157	61,652	66,235	71,140	76,387	123,986
Discount Rate	0.787033	0.619421	0.487505	0.383683	0.301971	0.237661	0.187047	0.147212	0.115861	0.091186
Present Value	5,650	25,701	24,088	20,393	17,259	14,652	12,389	10,472	8,850	11,305
(Total Present Value)										150,764
Internal Rate of Return (%)										27.0595

* Including Interest

X. Raw Materials for Nylon (Caprolactam)

1. Production Capacity and Site

1-1 Demand Forecast

Table X-1 shows a demand forecast on caprolactam in Indonesia. As shown evidently in Table X-1, the domestic demand for caprolactam is forecast to be 26,000 t/y in 1977, 41,000 t/y in 1979 and 51,000 t/y in 1981.

Table X-1 Demand Forecast on Caprolactam in Indonesia

(10^3 ton)

	Domestic Demand	Possible Export Amount to Southeast Asian Countries
1974	15	52
1975	18	40
1976	23	96
1977	26	166
1978	33	219
1979	41	263
1980	49	312
1981	51	350

1-2 Plant Scale

The operation to be undertaken in the caprolactam plant which will be constructed in Indonesia should be fully competitive internationally. Table X-2 shows a summary of capacity of caprolactam plants per unit production which have been constructed over the recent few years in the world in order to clarify the internationally competitive scale. As shown in Table X-2, the capacity per unit ranges approximately from 40,000 t/y to 60,000 t/y.

Table X-2 Caprolactam Plants Completed in the Recent Years

	Company	Process	Capacity (10^3 t/y)	Start Up Date
U.S.A.	Nypro	New DSM	50	1972
U.K.	Nypro	New DSM	50	1972
Belgium	BASF	BASF	60	1972
Italy	Chimica Dauna	SNIA	40	1973
U.S.S.R.	State Authority	DSM	50	1972
Mexico	Univex	DSM	44	1972
Japan	Toray	PNC	55	1971
Japan	Ube Ind.	New DSM	40	1972

1-3 Determination of Production Capacity

On the basis of the above-mentioned demand forecast and the recent plant scale, the following two cases can be considered as the production capacity figures for this project.

Case 1 : 60,000 t/y
Case 2 : 40,000 t/y

Case 1 pertains to a facility of maximum 60,000 t/y per unit which is the recent caprolactam plant capacity in the world, thereby giving an ample extent of international competitiveness. In this case, the domestic demand in Indonesia will be less than 60,000 t/y, so that the balance between the production and the domestic demand will be allocated to be export. The export ratio to the production will not be too excessive, i.e., 45% in 1978, 32% in 1979, 18% in 1980 and 15% in 1981.

Case 2 has the main objective of meeting the domestic demand, yet will not impede on the economical efficiency of the operation. In this case also the exportation will have to be carried out until 1978, however, in 1979 the production and the domestic demand will attain an equilibrium, so that, all the produced caprolactam will be consumed within Indonesia. Therefore, from 1980 onward, caprolactam will have to be imported in order to fulfill the domestic demand.

In order to produce 60,000 t/y of caprolactam, approximately 60,000 t/y of cyclohexane is necessary. If converted in terms of benzene, the required amount will be 56,000 t/y.

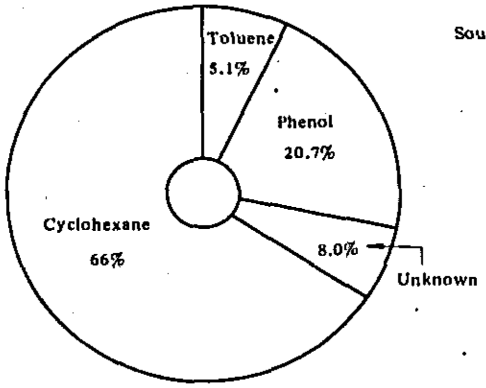
When a reformer of 10,000 to 15,000 BPSD is constructed for the production of BTX in Indonesia, approximately 100,000 t/y of benzene will be obtained. Therefore, there would be no problem in obtaining the raw materials for caprolactam production.

2. Process Outline

2-1 The Present Trend of the Process

The industrialization of caprolactam production has a comparatively long history. The production has so far been carried out by means of the so-called phenol process, in which the production is undertaken via phenol. However, in the recent years, a high purity cyclohexane has become available through low cost production.

On the other hand, along with the improvements made in the oxidation technique, the phenol process at present is being replaced by a new process in which the cyclohexane is employed as the starting raw material. Table V-56 shows the employed caprolactam production process and the relative production amount concerning various leading manufacturers of the world. Figure X-1 illustrates the ratio of the starting raw materials.



Source : Table V-56

Figure X-1 Starting Raw Material-wise Production Rate of Caprolactam

As is shown by this figure, 66% of the present caprolactam manufacturers of the world is employing cyclohexane as the starting raw material. The major reason is the advantage in the cost of the main raw material. The reason for the above is that the cyclohexane price is approximately 1/2 the phenol price and that the cyclohexane process is, therefore, superior in view of the production cost.

The classification of the presently employed processes can be shown as illustrated in Table X-3. At present, the highest extent of production is being carried out by the cyclohexane oxidation process followed by phenol process, the PNC process, the SNIA process, etc. All these processes have their own advantages and disadvantages and the industrial application of these are being made in accordance with the condition of each of the companies.

For the oxidation of cyclohexane, metallic catalysts such as naphthenic acid cobalt, etc. have conventionally been employed; however, in the recent years, boric acid catalysts are beginning to be employed in some cases in view of the more favorable rate of yield. Of these processes, the PNC process which applies photo-chemical reaction and the SNIA process which employes toluene as the starting raw material are attracting attention as being unique and outstanding.

Table X-3 Process-wise Classification of Caprolactam Production in the World

	10^3 t/y	%
Phenol Process	395	20.7
Direct Oxidation Process	1,116	58.4
DSM Process	392	(20.5)
BASF Process	380	(19.9)
Inventa Process	204	(10.7)
Bayer Process	140	(7.3)
PNC Process	150	7.8
SNIA Process	98	5.1
Others	152	8.0
Total	1,911	100.0

Source: Table V-56

As has been discussed in the foregoing, the caprolactam processes have been buttressed by a remarkable extent of growth in demand, and have so far been displaying developments of new processes, industrialization of the developed processes and the continuous replacement of the old by the new.

As a result, an outstanding pace of development in this field has been achieved so far. However, the problem which recently took place in this respect is the decrease of the by-produced ammonium sulfate. Originally, the production of caprolactam inevitably accompanies the by-production of ammonium sulfate the amount of which is higher than that of caprolactam which is the main product intended for the process (e.g., the by-production of ammonium sulfate sometimes amounts to 1.7 to 4.6 times the production amount of caprolactam). Such by-produced ammonium sulfate has so far been sold as fertilizer. However, when compared with urea, ammonium sulfate has approximately 1/2 of the nitrogen content. Also, for the most part, ammonium sulfate is not of the synthesis product but of by-produced or recovered products. This being the circumstance, ammonium sulfate cannot be sold on a high-price level.

Therefore, in the caprolactam production in which a large amount of ammonium sulfate is by-produced, the price of the ammonium sulfate will directly reflect upon the production cost of caprolactam. This being the circumstance, the amount of the by-produced ammonium sulfate at present has come to assume an important problem within the framework of caprolactam production process. In order to cope effectively with this problem, a consecutive development of new processes has been undertaken in which the major objective is to reduce the extent of the ammonium sulfate by-production.

These newly developed processes are being placed under actual operation. From this viewpoint, the New DSM process and PNC process can be named as the ones which are assuming the most advantageous position at the present stage. The details of these processes will be explained in the latter part of this writing.

As has been discussed in the foregoing, the problems pertaining to the by-produced ammonium sulfate will increase their seriousness year aftet year in the future and the general trend will be the development of even newer processes in order to solve these problems.

2-2 Outline of Caprolactam Production Processes

The caprolactam production has been industrialized mainly on the basis of the processes in which phenol, cyclohexane, toluene, etc. are being employed as the starting raw materials. Figure X-2 shows the comparison of these processes. Figure X-3 shows the nitroso group introduction process to the cyclohexane in these processes. The caprolactam production processes have been developed by means of combining these two reactions together with the incorporation of the know-how possessed by each one of the manufacturers. Particularly, in the case of the cyclohexane oxidation process, there are a number of variations and they are individually called after the names of companies who employ them.

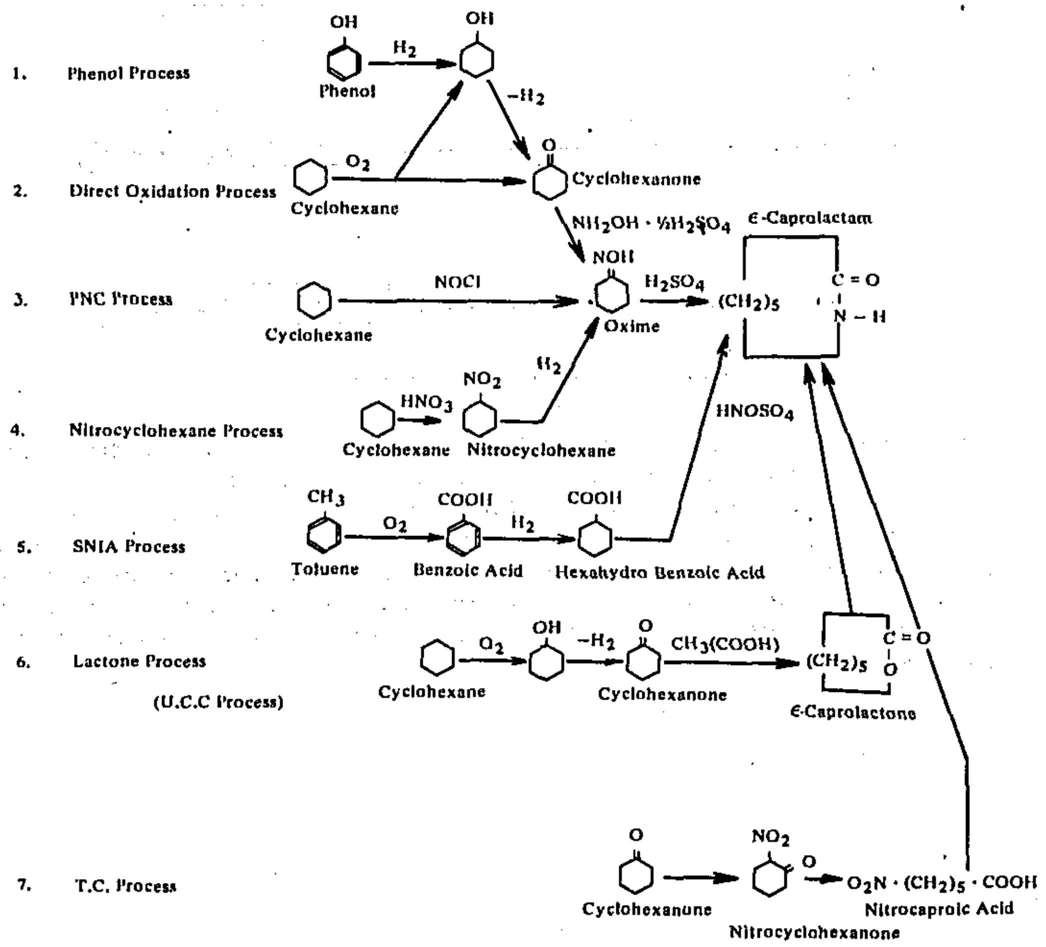


Figure X-2 Major Reaction Process of Caprolactam

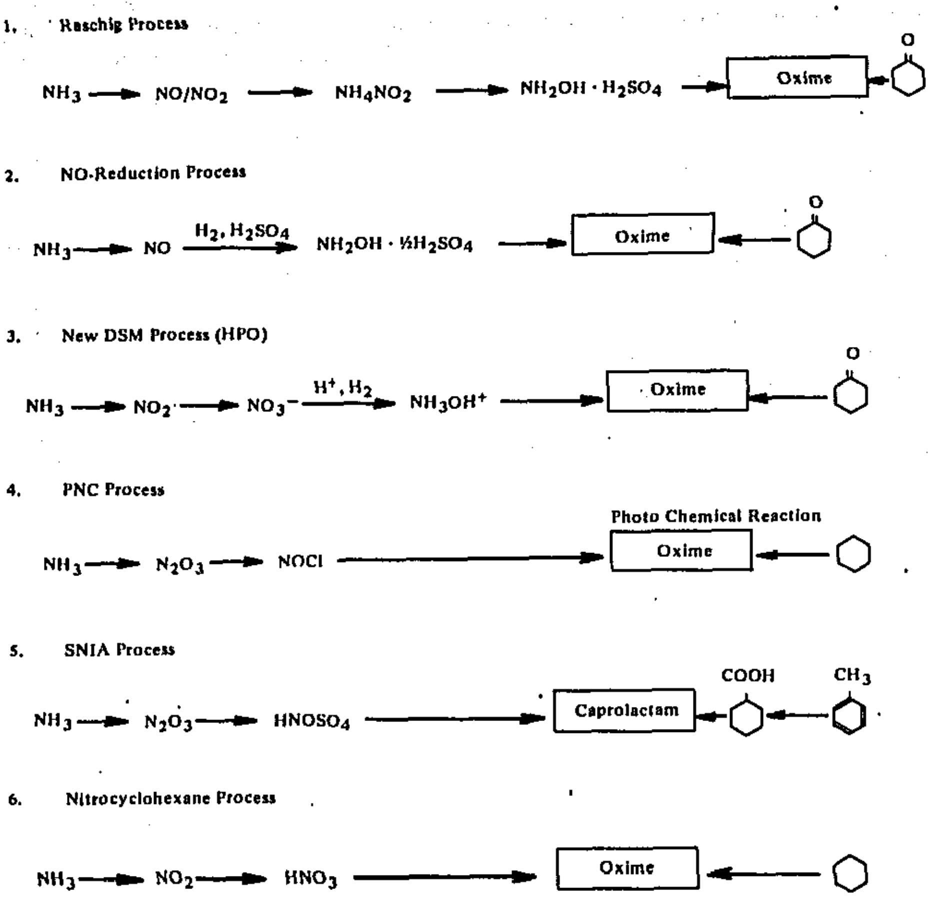


Figure X-3 Nitroso Group Introduction Process

As is shown in Figure X-2 and X-3, there are a number of different processes and the combination of these will make considerable amount of variations. In this respect, it is considered that each manufacturer is carrying out the operation by combining the optimum processes in view of the requirements of the plant site conditions, production scale, the own history of process development, etc. For the description of the outline of each one of these, Figure X-4 illustrates the major combinations of the reaction processes concerning each of the processes.

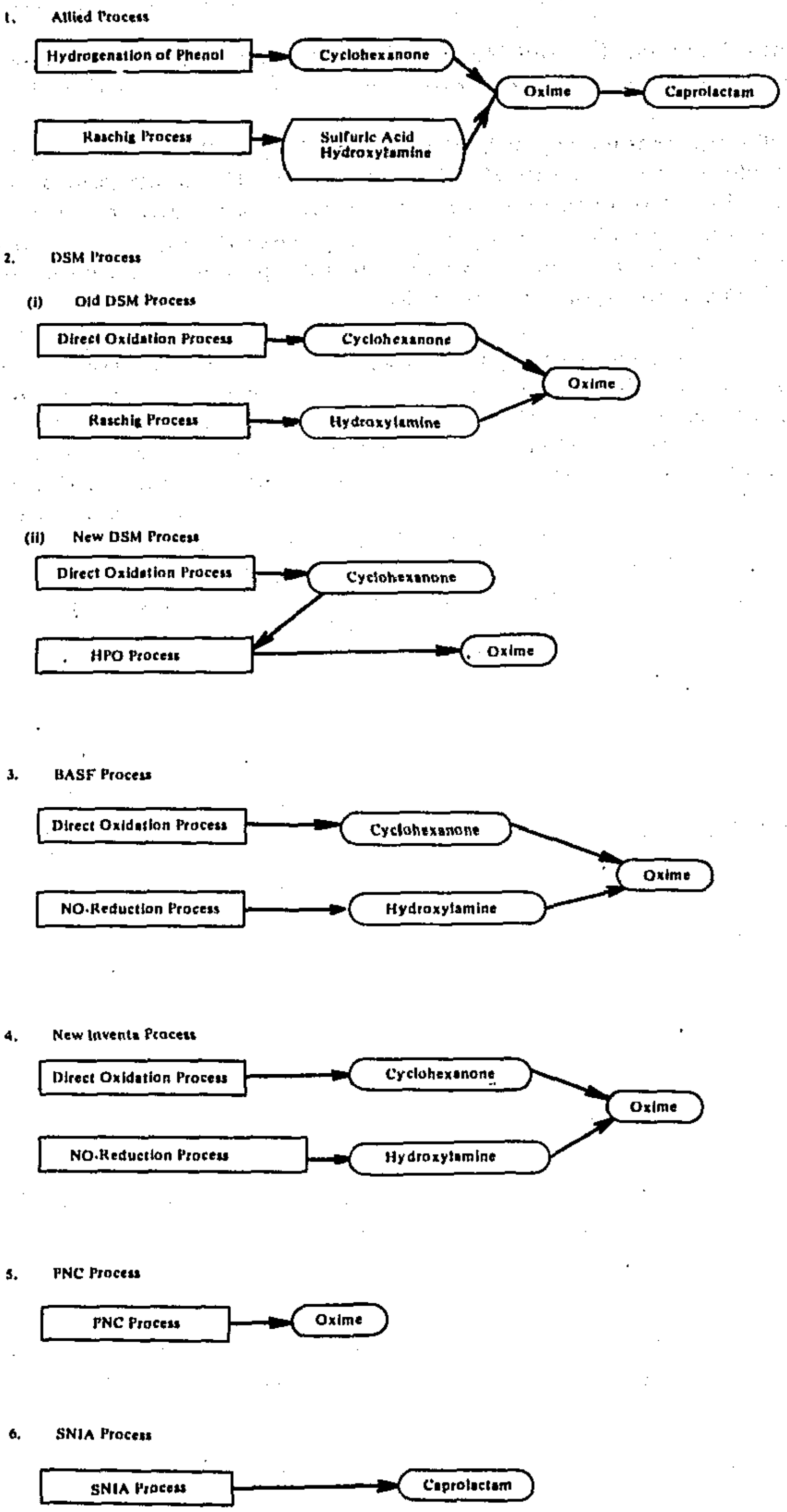


Figure X-4 Combination of Reaction Processes for Caprolactam Production

2-3 Outline of Various Processes and Comparison Thereof

2-3-1 Phenol Process (Allied Process and Raschig Process Hydroxylamine Production)

The phenol process was first developed by Allied Chemical and the industrial production by means of this process is still continued since 1954. In this process, phenol is employed as the starting raw material in order to carry out the caprolactam production. Even at the present stage, it is said that the various new developments centering around this process are being undertaken in order to enhance the quality and yield.

In a series of reaction equipment phenol is catalytically hydrogenized and is converted into cyclohexanone and minor quantities of cyclohexanol. The product stream from the reactor contains un-reacted phenol which is removed by distillation. Also, the catalyst and recovered phenol are recycled with fresh phenol feed. The mixture of the cyclohexanone and cyclohexanol are separated by distillation. The recovered cyclohexanone is fed into the oximation process. The Allied process uses the Raschig process for the production of hydroxylamine. The outline of the Allied process is shown in Figure X-5. Hydroxylamine sulfate production by the Raschig process is as follows. The Raschig process uses sulfur dioxide, ammonia, carbon dioxide and water as the raw materials.

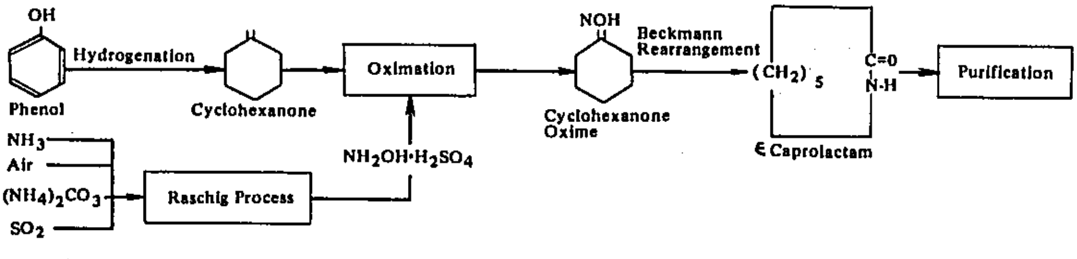


Figure X-5 Process Block Diagram of Allied Process

Ammonium carbonate and ammonium bicarbonate are produced by absorbing ammonia and carbon dioxide in water.

$$3NH_3 + 2CO_2 + 2H_2O \longrightarrow (NH_4)_2CO_3 + NH_4HCO_3 \quad \cdots\cdots\cdots(1)$$

Ammonium nitrite is produced by catalitically oxidizing ammonia into NO/NO_2 and absorbing cooled NO/NO_2 in ammonium carbonate solution while generating carbon dioxide.

$$2NH_3 + 3O_2 \longrightarrow NO \cdot NO_2 + 3H_2O \quad \cdots\cdots\cdots(2)$$

$$NO \cdot NO_2 + (NH_4)_2CO_3 \longrightarrow 2NH_4NO_2 + CO_2 \quad \cdots\cdots\cdots(3)$$

Hydroxylamine disulfonate is produced by absorbing sulfur dioxide in ammonium nitrite.

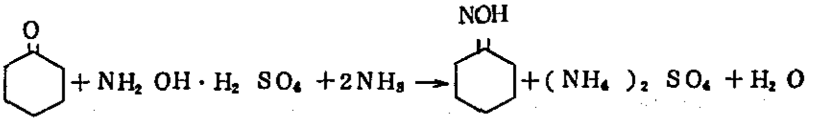
$$NH_4NO_2 + NH_3 + 2SO_2 + H_2O \longrightarrow HON(SO_3NH_4)_2 \cdots\cdots(4)$$

The hydroxylamine disulfonate is hydrolyzed to hydroxylamine sulfate and ammonium sulfate, then the products are separated.

$$HON(SO_3NH_4)_2 + 2H_2O \longrightarrow NH_2OH \cdot H_2SO_4 + (NH_4)_2SO_4 \cdots(5)$$

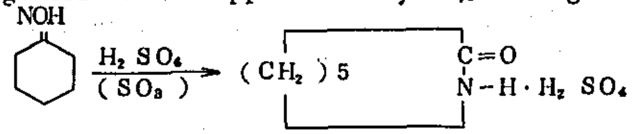
Hydroxylamine sulfate is fed to the oximation process, while ammonium sulfate is fed to the recovery plant and high quality ammonium sulfate free from organic substances is recovered.

Hydroxylamine sulfate is reacted with cyclohexanone and ammonia on the following oximation process, to produce cyclohexanone oxime. At this time, ammonium sulfate of mol equivalent to oxime is by-produced.



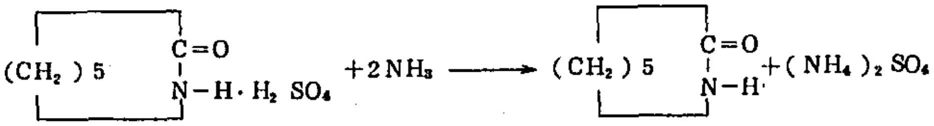
……(6)

The separated cyclohexanone oxime is fed to the Beckmann rearrangement vessel together with an approximately 20% fuming sulfuric acid.



……(7)

The crude caprolactam which consists of caprolactam and sulfuric acid are added by ammonia, water, and solvent then, fed to the neutralizer and neutralized by sulfuric acid and at the same time, the caprolactam is extracted from the ammonium sulfate solution.



··(8)

The products obtained from the neutralization process is treated in a separator in which the solvent/caprolactam layer is separated from the ammonium sulfate solution layer.

The ammonium sulfate layer is extracted by solvent and the remaining caprolactam is removed. From the ammonium sulfate solution, the solvent is removed by gas phase separation and, treated ammonium sulfate solution is fed to the ammonium sulfate unit. The solvent/caprolactam layer is sent to the distillation tower where the solvent is separated from the caprolactam layer. The caprolactam is then fed into the crude caprolactam tank from where it is further sent to the purification process.

2-mol of ammonium sulfate is by-produced per 1-mol of cyclohexanone oxime at the hydroxylamine producing steps (Formula 5) and at the oxime producing steps (Formula 6) and further, 1.5 to 2-mol of ammonium sulfate is by-produced at the neutralization step (Formula 8) after the Beckmann rearrangement. Therefore, the maximum 4.6 kgs of ammonium sulfate is by-produced per 1 kg of caprolactam.

Caprolactam is purified by crystallization. The crude caprolactam proceeds in one direction with the recycle mother liquor passing counter-currently to the stream of crystals moving forward to be purified. The crystal caprolactam from the crystallization process is melted, dried and flaked. Ammonium sulfate is crystallized by evaporating water in crystallizer. Thus produced crystals are dehydrated by a centrifuge and dried to give a final product.

In the foregoing paragraphs, the outline of the Allied process was described. As this Allied process employs phenol as the starting raw material and because of the high cost level of phenol, the raw material cost level of this process is consequently higher when compared with other processes which employ cyclohexane as the starting raw material. This being the case, it is suspected that the Allied process has already lost its competitiveness.

Due to the fact that the Raschig process is employed in the Allied process, the amount of the by-produced ammonium sulfate is approximately 4.6 kg for 1 kg of cyclohexanone oxime as has already been discussed. Compared with the other processes (PNC process, New DSM process and BASF process), this level of by-production is on a considerably higher side, thereby creating a disadvantage in the cost.

In each of the process stages of the whole process, special materials for the equipment are employed. However, such a necessity would not present any serious problem in view of the total scale of the plant. By means of employing this process, it is possible to construct a continuous single-pass production line of a capacity up to 90,000 t/y. Table X-4 shows the raw material unit consumption and utilities unit consumption in the case of the Allied process.

Table X-4 Unit Consumption of Raw Materials and Utilities in the Allied Process

		Unit Consumption (kg/kg)
Raw Materials	Phenol	0.89
	Hydrogen	0.04
	Fuming Sulfuric Acid	1.35
	Sulfuric Acid	0.68
	Ammonia	1.46
	Carbon Dioxide	0.52
	Ammonium Sulfate	(-)4.6
Utilities	Electricity	0.9 kwh/kg
	Steam	11
	Cooling Water	1.7 ton/kg

Source: European Chemical News May 2, 1969

2-3-2 DSM Process

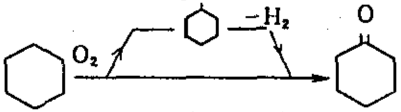
DSM commenced commercial production in 1967 after constructing a plant by employing the cyclohexane oxidation process which has been developed by DSM themselves.

For the production of hydroxylamine sulfate, the conventionally employed process has been the Raschig process. However, in 1965, a new process for hydroxylamine production was developed in which no ammonium sulfate by-production took place. The newly developed process is being employed for actual operations in some of the plants. This being the circumstance, the following two processes are concurrently employed as far as the DSM process is concerned.

(1) Old DSM Process : Cyclohexane oxidation process (plus the Raschig process)

(2) New DSM Process : Cyclohexane oxidation process (plus the HPO process)

This process employs the liquid-phase air-oxidation process of cyclohexane. Cyclohexanone is produced in two stages, i.e., in the first stage, the cyclohexane is converted by air into cyclohexanone and cyclohexanol; in the second stage, the cyclohexanol is converted into cyclohexanone by catalytic dehydrogenation.



Originally, DSM employed a cobalt catalyst for the reaction. However, due to the necessity of further improvement, an introduction of the reactor design technique from IFP (Institut Francais du Petrole) in which boric acid is employed as the catalyst. The advantage of this process is the fact that the consumption extent of cyclohexane and caustic soda will be less, thereby further reducing the production cost. At first the DSM process, both the old and the new, employ cyclohexane as the starting raw material because of the employment of the cyclohexane oxidation process. Therefore, it is obvious that this process has a greater advantage in the cost aspect over the phenol process.

However, in view of the facilities installation aspect, the reaction conversion rate per one pass is approximately 5% to 15% due to the necessity for minimizing, as much as possible, the formation of by-product at the stage of oxidation of cyclohexane. This necessity is present commonly for all types of cyclohexane oxidation process.

This implies that a large amount of recycle stream of cyclohexane into the reactor is made, thereby compelling the refining facilities becoming large in size. At the same time, it is expected that the utilities unit consumption are on a considerably higher level concerning such items as the heating steam required at the time of recovering.

Regarding the ammonium sulfate, there is no conspicuous difference between the Old DSM process which employs the Raschig process and the Allied process (4.5 tons ammonium sulfate by-production/ton of caprolactam). On the other hand, the new hydroxylamine process developed by DSM does not by-produce ammonium sulfate at all up till the cyclohexanone oxime producing stage. This new process (hereinafter referred to as the HPO process) will be described briefly in the following paragraphs.

This new process can be divided into four stages:

(1) Production of hydroxylamine by hydrogenation of nitrate ions over a noble metal catalyst.

(2) Reaction of hydroxylamine and cyclohexanone in the presence of a solvent.

(3) Separation of the solvent from oxime.

(4) Simultaneous decomposition of the ammonium ion contained in the circulating inorganic liquid and then NO/NO_2 is absorbed in the liquid.

The production of hydroxylamine is carried out in a gas/liquid contact reactor in which hydrogen gas is brought in contact with inorganic circulating liquid containing nitrate ions, buffer acid and precious metal catalysts. Through this reaction, the nitrate ions are hydrogenated, thereby producing hydroxylamine.

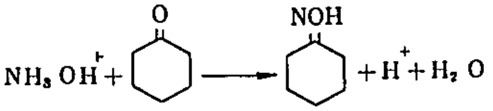
$$NO_3^- + 2H^+ + 3H_2 \longrightarrow NH_3OH^+ + H_2O$$

Buffer Solution : $HX \longrightarrow X^- + H^+$

At this time a small extent of ammonium ions are by-produced.

$$NO_2^- + 2H^+ + 4H_2 \longrightarrow NH_4^+ + H_2O$$

This ammonium ions do not interfere with the reaction, and thereafter be destroyed. At the second stage, hydroxylamine react with pure cyclohexanone in the presence of catalysts, thereby extracting oxime. Thus, the separation of the inorganic water solution and dissolved oxime is carried out.



Regarding the process stages after oxime production, the DSM process are almost identical to that of Allied and other processes.

The caprolactam production process developed by DSM into which the new hydroxylamine production (the New DSM process) is incorporated entails to the by-production of ammonium sulfate from the sulfuric acid neutralization process after the Beckmann rearrangement, and, the by-production extent is 1.7 tons as against 1 ton of caprolactam which is extremely low when compared with the Allied process.

This New DSM process was employed by Nypro (U.K.) and by Ube Ind. (Japan). However, it is reported that problems in operation took place at the operation stage. In view of the scale of a plant, if it is desired to employ only one distillation tower for a cyclohexane oxidation plant of more than 35,000 tons of annual production capacity, the size of the tower will be exceedingly large so that the transportation of the distillation tower at the time of plant construction will become impossible. Therefore, in order to accommodate this level of production capacity, it is recommended that two distillation towers be employed. Tables X-5 and X-6 show the raw materials unit consumption and the utilities unit consumption in the Old and New DSM processes.

Table X-5 Unit Consumption of Raw Materials and Utilities in the Old DSM Process

		Unit Consumption (kg/kg)
Raw Materials	Cyclohexane	1.06
	Ammonia	1.51
	Caustic Soda	0.114
	Sulfur Dioxide	1.32
	Carbon Dioxide	0.41
	Fuming Sulfuric Acid	1.34
	Ammonium Sulfate	(-)4.5
Utilities	Electricity	0.46 kwh/kg
	Steam	13.1
	Cooling Water	1.65 ton/kg
	Process Water	0.010 ton/kg

Source: European Chemical News May 2, 1969

Table X-6 Unit Consumption of Raw Materials and Utilities in the New DSM Process

			(/kg Caprolactam)
Raw Materials	Benzene		0.995 kg
	Cyclohexane	0.905 kg	
	Ammonia	0.805 kg	0.743 kg
	Sulfuric Acid	1.368 kg	1.36 kg
	Hydrogen	0.096 kg	1.814 m^3
	Ammonium Sulfate	-1.8 kg	-1.8 kg
Utilities	Electricity	0.224 kwh	0.45 kwh
	Steam	6.60 kg	12.65 kg
	Cooling Water	1.10 m^3	1.62 m^3
	Boiler Feed Water	0.0025 m^3	0.005 m^3
	Fuel		470 kcal

Source: Hydrocarbon Processing, Nov. 1973, p.110
Hydrocarbon Processing, Nov. 1972, p.92 - 94

2-3-3 BASF Process

The BASF process originally employed phenol as the starting raw material. However, along with the development of the cyclohexane oxidation process, they have already given up the employment of the phenol process and have totally shifted to the cyclohexane oxidation process.

In the case of BASF the cyclohexane oxidation process is of a multiple-stage air oxidation process and improvements have been made in the refining process. The feature of the BASF process lies in the production process of hydroxylamine for which the NO-Reduction process is employed.

In the Raschig process, N_2O_3 is obtained by means of the oxidation of ammonia, however, in the NO-Reduction process, the oxidation is carried out not by air but by pure oxygen in order to obtain NO, as shown in the following formulae.

$$4NH_3 + 5O_2 \longrightarrow 4NO + 6H_2O \quad \cdots\cdots\cdots\cdots(1)$$

$$NO + \frac{3}{2}H_2 + \frac{1}{2}H_2SO_4 \longrightarrow NH_2OH \cdot \frac{1}{2}H_2SO_4 \quad \cdots\cdots(2)$$

Owing to this improvement, the hydroxylamine production process has been greatly simplified as compared with the conventional process. Further, it is reported that a drastic reduction in the amount of by-produced ammonium sulfate will be achieved. Theoretically speaking, the amount of ammonium sulfate by-produced at the oximation process will be reduced by 1/2 of that in the case of the Raschig process.

The shortcoming of this process is, although it is a common problem for all types of cyclohexane oxidation processes, the 1-pass yield of cyclohexane oxidation must be reduced. Otherwise, the amount of by-products will increase. Because of this fact, the increment in the plant cost and in the utilities unit consumption will be inevitable.

Table X-7 shows the raw materials unit consumption in the BASF process.

2-3-4 Inventa Process

The Inventa process is one of the cyclohexane oxidation processes employing cyclohexane as the raw material. Further, for the oxidation of cyclohexane, boric acid catalyst is not employed and instead, metallic catalysts are used. For the production of hydroxylamine, the Raschig process was employed at the early stage, however, the introduction of the NO-Reduction process have since been carried out, thereby reducing the extent of ammonium sulfate by-production. Although the thin layer distillation process has been employed conventionally for the purification of caprolactam, the operation has recently been switched over to the melt crystallization process.

Table X-7 **Unit Consumption of Raw Materials in the BASF Process**

		Unit Consumption (kg/kg)
Raw Materials	Cyclohexane	1.150
	Ammonia	0.953
	Sulfuric Acid	0.557
	Fuming Sulfuric Acid	1.370
	Caustic Soda	-
	Hydrogen	0.467 m^3
	Oxygen	0.559 m3
	Ammonium Sulfate	-2.7

Source: Kagaku Koogyo (Chemical Industry) No. 10, p. 62 - p. 71 (1970)

When compared with the DSM process, each one of the cyclohexane oxidation processes presents individual differences in reaction conditions. However, it is assumed that, basically all the processes in this category are highly analogous. Also, the production method of hydroxylamine is, as has been mentioned earlier, of the NO-Reduction process which is almost identical to the BASF process. Therefore, when compared with the BASF process, the NO-Reduction process has no conspicuous difference in view of its features and shortcomings.

Lately, Inventa developed an improved process of cyclohexane oxidation process, i.e., the process which is popularly called the New Inventa process. It is said that the actual operation by employing this newly developed process has already started. The feature of this process is the reduction of the oxygen concentration at the time of air blow-in in order to enhance the yield of cyclohexanone. Table X-8 shows the raw materials unit consumption and utilities unit concumption in the case of the New Inventa process.

Table X-8 Unit Consumption of Raw Materials and Utilities in the New Inventa Process

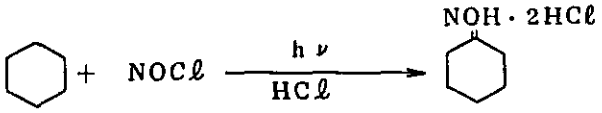
		Unit Consumption (kg/kg)
Raw Materials	Cyclohexane	1.01
	Caustic Soda	-
	Ammonia	0.96
	Sulfuric Acid	0.7
	Fuming Sulfuric Acid (SO_3 100%)	1.1
	Oxygen	0.38 m^3/kg
	Hydrogen	0.47 m^3/kg
	Ammonium Sulfate	-2.6
Utilities	Electricity	0.9 kwh/kg
	Steam	12
	Fuel Oil	3.0 x 10^3 kcal/kg
	Cooling Water	0.14 ton/kg

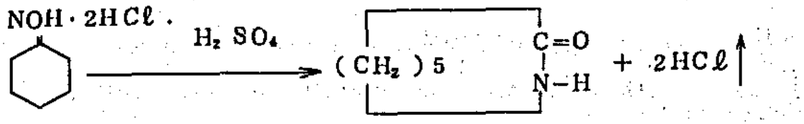
Source: Hydrocarbon Processing, Nov. 1973
European Chemical News, May 2, 1969

2-3-5 PNC Process

This is a process developed by Toray having a feature of producing cyclohexanone oxime by directly introducing nitroso group into cyclohexane by employing photo-chemical reactions. Because of this provision, the reaction process is the shortest of all the other processes. While the conventional processes required four-stage reaction processes when employing phenol or cyclohexane as the starting raw material, the PNC process reduced the reactions into only two. Also, due to the unnecessity in this process for the production of cyclohexanone by the oxidation of cyclohexane, the required amount of cyclohexane for the production of caprolactam is reduced.

Further, while in the other processes comparatively high-cost hydroxylamine is utilized as the nitrogen source, the PNC process employs as the nitrogen source comparatively low-cost nitrosyl chloride. Due to this feature, no ammonium sulfate by-production takes place at this step. Therefore, the by-production amount of ammonium sulfate against one ton of caprolactam produced is only 1.7 tons which is low when compared the rest of the processes. The following shows the formula of the photo-reactions and the rearrangement-reactions.





In the PNC process, the shortcoming is the necessity for employing special materials to build the reaction equipment because of the handling of highly corrosive gas and liquid such as hydrogen chloride, nitrosyl chloride and mixture of these chemicals and cyclohexane. Table X-9 shows the raw materials unit consumption and utilities unit consumption in the case of the PNC process.

Table X-9 Unit Consumption of Raw Materials and Utilities in the PNC Process

		(/kg Caprolactam)
Raw Materials	Cyclohexane	0.908 kg
	Ammonia	0.649 kg
	Sulfuric Acid Fuming Sulfuric Acid	1.279 kg
	Ammonium Sulfate	-1.70 kg
Utility	Electricity	4.050 kwh

Source: Process Handbook 1973
(Japan Petroleum Institute)
European Chemical News, Jan. 26, 1968

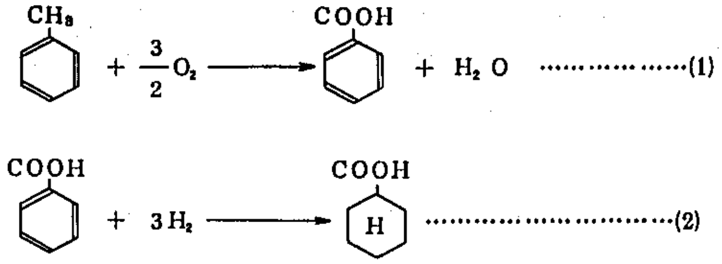
2-3-6 SNIA Process

This process was developed by Snia Viscosa of Italy. The feature of this process is the employment of toluene, which is lower in cost than benzene, as the starting raw material.

In the SNIA process, hydrogenation of toluene is undertaken in order to convert it into hexahydrobenzoic acid which is reacted with nitrosyl sulfuric acid in the presence of fuming sulfuric acid in order to directly produce caprolactam. This being the case, the production reactions do not go through the cyclohexanone oxime.

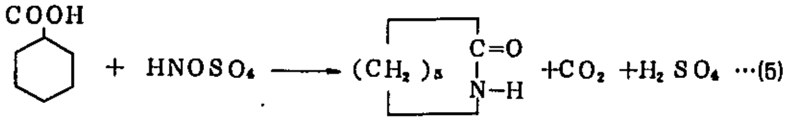
The main raw materials used in this process are; air, toluene, hydrogen, ammonia and sulfuric acid. Nitrosyl sulfuric acid is employed as the nitrogen source to be introduced into the cyclohexane. The by-produced ammonium sulfate is slightly lower in amount when compared with the air oxidation of other processes, however, the difference is not significant.

The main reactions of this process can be expressed by means of the following formulae:



$2NH_3 + 3O_2 \longrightarrow N_2O_3 + 3H_2O$ ························(3)

$H_2SO_4 + N_2O_3 + SO_2 \longrightarrow 2HNOSO_4$ ···············(4)



Benzoic acid can be produced with a yield rate of 93% to 94% (if toluene is air-oxidized at 150 to 170°C in the presence of a cobalt catalyst). This reaction is expressed in the above formula (1). Benzoic acid is hydrogenated quantitatively to hexahydrobenzoic acid by using a 5% palladium·carbon catalyst at 150°C, 1 atm. This reaction is expressed in the above formula (2). As shown by formula (5), hexahydrobenzoic acid is reacted with nitrosyl sulfuric acid in the presence of fuming sulfuric acid at 80°C, the nitrosation reaction is carried out while generating carbon dioxide, and simultaneously Beckmann rearrangement takes place, thereby producing caprolactam.

The yield of caprolactam from hexahydrobenzoic acid is 95% to 96% (90% on nitrogen basis). The reaction temperature is controlled by the evaporation heat of cyclohexane. The reactor effluent is hydrolized and the unconverted hexahydrobenzoic acid is extracted by cyclohexane. The caprolactam solution containing less than 0.1% of hexahydrobenzoic acid is neutralized by ammonia. The caprolactam is extracted from the organic phase by aromatics or chlorinated hydrocarbon solvent, then washed and extracted by water. The extracted caprolactam is purified by distillation.

Concerning the cost of caprolactam, no significant difference exists in this process when compared with the cyclohexane oxidation processes, etc., because of the fact that the cost of sub-raw materials and chemicals are high although the unit cost of toluene is low. The shortcoming in this process is the high extent of ammonium sulfate by-production, i.e., 4.1 ton per ton of caprolactam.

The raw materials unit consumption in the case of the SNIA process is shown in Table X-10.

Table X-10 Unit Consumption of Raw Materials in the SNIA Process

		Unit Consumption (kg/kg)
Raw Materials	Toluene	1.11
	Cyclohexane	0.03
	Hydrogen	0.08
	Fuming Sulfuric Acid	2.90
	Ammonia	1.25
	Caustic Soda	0.16
	Ammonium Sulfate	(-)4.1

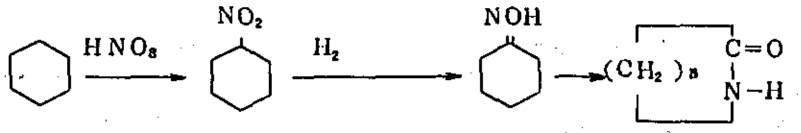
Source: Hydrocarbon Processing, Nov. 1973
European Chemical News, July 11, 1969

2-3-7 Nitrocyclohexane Process

This nitrocyclohexane process has been developed by Du Pont and carries out the nitrosation of cyclohexane. The obtained nitrocyclohexane is reduced to produce cyclohexanone oxime.

The reactions in this process are carried out in the liquid phase or gas phase by using nitric acid or nitrogen peroxide as nitrosation agent. The nitrosation is carried out at 120 to 125°C in liquid phase process at 3 to 5 atm, by 35% nitric acid. 50% of adipic acid to nitrocyclohexane is by-produced. After being washed by caustic soda, nitrocyclohexane is recovered by distillation. Nitrocyclohexane is reduced by hydrogen in the presence of zinc, chrome, silver oxide catalyst at 140 to 160°C and 100 to 200 atm, to convert nitrocyclohexane into cyclohexanone oxime. The yield is 80% to 90% and the by-product are cyclohexylamine or cyclohexyl-hydroxylamine, etc.

The oxime is extracted by a solvent and converted into caprolactam in accordance with the conventional process.



The advantage of this process is the comparative shortness of the process itself. However, the problems are the danger involved in the operation of the nitrosation process and the considerable amount of by production of such as adipic acid and cyclohexylamine. No plant is now being operated by employing this process for the commercial production of caprolactam.

2-3-8 Techni-Chem Process

The Techni-chem process for producing caprolactam from cyclohexanone has the advantage that the extent of the by-product is extremely small when compared with other processes, although the commercial production of this process has not yet been carried out.

In this process, the production of caprolactam is conducted after going through an extremely long process involving cyclohexane, cyclohexenyl acetate, nitrocyclohexanone, nitrocaproic acid and ε-amino caproic acid. Although this process has an attractive feature of by-producing no ammonium sulfate, the process from the cyclohexane up to the production of caprolactam is highly lengthy, therefore, economics of this process will be inferior.

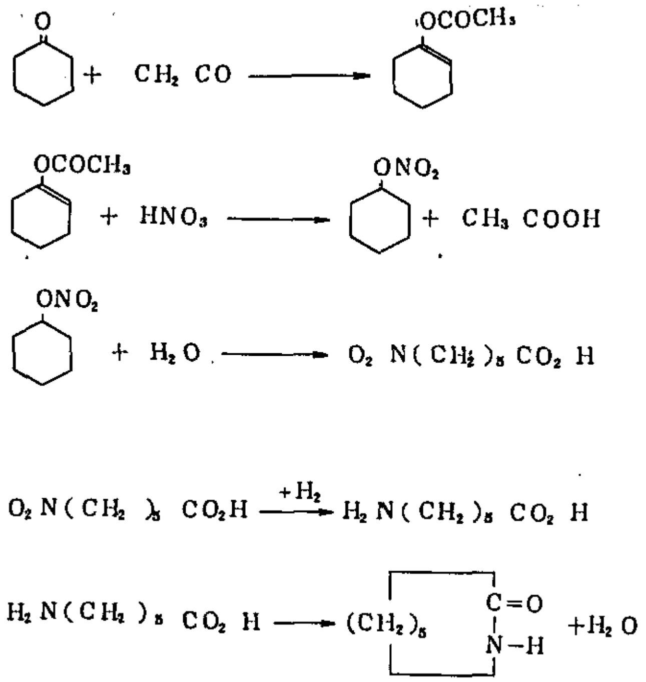


Table X-11 shows the raw materials unit consumption and utilities unit consumption in the case of the Techni-chem process.

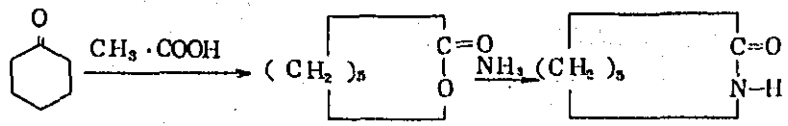
Table X-11 Unit Consumption of Raw Materials and Utilities in the Techni-Chem Process

		Unit Consumption (kg/kg)
Raw Materials	Cyclohexanone	1.0
	Hydrogen	0.07
	Acetic Acid	0.1
	Nitric Acid	0.62
Utilities	Electricity	0.28 KWH/kg
	Natural Gas	3.0×10^3 kcal/kg
	Stream	5.7
	Cooling Water	$0.48\ m^3/kg$

Source: European Chemical News, May 2, 1969

2-3-9 Lactone Process (U.C.C. Process)

This process was developed by Union Carbide Company in the U.S.A. The Lactone process carries out a further oxidation of cyclohexanone which is obtained by oxidation of cyclohexane by using a peracetic acid into caprolactone, and then caprolactone is reacted with ammonia to produce caprolactam.



This process by-produces 0.75 ton to 1 ton of acetic acid for 1 ton of caprolactam. This process by-produces a large quantity of acetic acid, although it does not by-produce ammonium sulfate. Union Carbide constructed a plant by employing this process in 1970, however, due to technical problems the company is reported to have abandoned the production of caprolactam and switched the operation to the production of caprolactone.

2-3-10 Process Selection

Table X-12 shows the outline of the above-mentioned production processes. The following are the comparison of various processes.

(1) Actual Records of Commercial Production

The following processes have already accomplished commercial production. Processes other than the following cannot be considered as the subjects for the selection for the industrialization project in Indonesia.

(a) Allied Process: For the production of cyclohexane from phenol and the production of hydroxylamine by the Raschig process.

(b) Old DSM Process: The cyclohexane oxidation process and the Raschig process

(c) New DSM Process: The cyclohexane oxidation process and the HPO process

(d) BASF Process: The cyclohexane oxidation process and the NO-Reduction process

(e) New Inventa Process: The cyclohexane oxidation process and the NO-Reduction process

(f) PNC Process:

(g) SNIA Process:

Table X-12 Comparison of Various Caprolactam Production Processes (I)

	PNC	SNIA	Nitrocyclohexane	Techni-Chem	Lactone
Actual Records of Production (10^3 t/y)	150	98	NO Production at Present	NO Production	NO Production
Actual Records of Production (%)	(7.8)	(5.1)			
Main Raw Materials, Process	Photo-nitrosation of cyclohexane by nitrosyl chloride	Toluene, benzoic acid, hexahydrobenzoic acid reaction is undertaken and then caprolactam production is carried out by reaction with nitrosyl sulfuric acid.	Nitro-cyclohexane is produced from cyclohexane and caprolactam production is carried out via cyclohexanone oxim.	ϵ-amino caproic acid is produced through a number of steps from cyclohexane in order to produce caprolactam.	Caprolactam production is carried out through cyclohexane, cyclohexanone. caprolactam.
Hydroxylamine Production Process	-	-	-	-	-
Sub-raw Materials & Others (ton/ton caprolactam)	Nitrosyl chloride production by a reaction of nitrosyl sulfuric acid and hydrogen chloride	Nitrosyl sulfuric acid production by utilizing ammonia and sulfuric acid as raw materials	-	Employment of Keten	Employment of peracetic acid
By-produced Ammonium Sulfate (ton/ton caprolactam)	1.7	4.1	?	0	(Acetic Acid 0.75)
Remarks	The number of reaction steps has been drastically reduced by the employment of photo-chemical reaction. Due to the absence of cyclohexane oxidation step, higher yeild is available. The ammonium sulfate by-production is also low as in the case of New DSM Process. However, special materials and equipment must be employed and the electrical power consumption is high.	Due to employment of toluene as the raw material, raw material cost reduction can be expected. The amount of ammonium sulfate by-production is slightly lower than a number of phenol process and Old DSM Process.	This process is not actually operated at present. The shortcoming is the high amount of by-production of adipic acid and cyclohexylamine.	No actual industrialization record is achieved. The steps from cyclohexane up to caprolactam are extremely long.	No industrial production record has been achieved. Although no ammonium sulfate is by-produced, the acetic acid by-production takes place.

Table X-12 Comparison of Various Caprolaction Production Processes (2)

	Phenol	Old DSM	New DSM	BASF	New Inventa
Actual Records of Production (10^3 t/y)	395	392	392	380	204 1)
Actual Records of Production (%)	(20.7)	(20.5)	(20.5)	(19.9)	(10.7) 1)
Main Raw Materials, Process	Hydrogenation of Phenol, Dehydrogenation	Direct Oxidation of Cyclohexane (Co Catalyst)	Direct Oxidation of Cyclohexane (Co or Boric Acid-Catalyst)	Direct Oxidation of Cyclohexane (Co Catalyst)	Direct Oxidation of Cyclohexane (Co Catalyst)
Hydroxylamine Production Process	Raschig	Raschig	HPO	NO-Reduction	NO-Reduction
Sub-raw Materials & Others (ton/ton caprolactam)	-	-	-	-	-
By-produced Ammonium Sulfate (ton/ton caprolactam)	4.6	4.5	1.8	2.7	2.6
Remarks	Yield is considered to be high and refining and reactions can be easily undertaken because of employment of phenol as the raw material. However, due to the employment of phenol and the Raschig Process, cost increment is high. The ammonium sulfate by-production is high.	Although popularly employed internationally, the process is complicated and a ammonium sulfate by-production is high due to the employment of the Raschig Process.	This is a process in which the hydroxylamine production process is improved, thereby drasticaly decreasing the ammonium sulfate by-production. However, the actual records is poor at present and technological improvements seem necessary for a stable operation.	This is a process popularly employed in the world. Improvements were made in the hydroxylamine production process thereby reducing the ammonium sulfate by-production. However, the amount of ammonium sulfate by-production is considerably higher than New DSM Process and PNC Process.	This is a process popularly employed in the world. Improvements were made in the hydroxylamine production process thereby reducing the ammonium sulfate by-production. However, the amount of ammonium sulfate by-production is considerably higher than New DSM Process and PNC Process.

Note: 1) Including Old Inventa Process

(2) Starting Raw Materials

The production of caprolactam can be classified into three categories according to the starting raw materials, i.e., phenol in the Allied process, toluene in the SNIA process and cyclohexane in other processes. More than 70% of caprolactam in the world is produced by processes which utilize cyclohexane as the raw material. As far as the raw material cost is concerned, it is naturally forecast that this tendency will persist for some time to come, so that the following four processes and the Old DSM process can be considered for the selection of the caprolactam production process. However, as the New DSM process has been developed, the Old DSM process will be excluded from the scope of selection.

1. New DSM Process
2. BASF Process
3. New Inventa Process
4. PNC Process

(3) By-Produced Ammonium Sulfate

A serious problem in all the caprolactam processes is the ammonium sulfate by-production. As mentioned earlier, the cost of caprolactam is affected by the by-produced ammonium sulfate. Therefore, all the companies give priority to the development of, and the replacement with, a new process which decreases the quantity of the by-produced ammonium sulfate. Of the above-mentioned processes, the following are the four processes enumerated in the increasing order of the amount of by-produced ammonium sulfate.

PNC Process	1.7 (ton	ammonium sulfate/ton	caprolactam)	
New DSM Process	1.8 (	"	"	)
New Inventa Process	2.6 (	"	"	)
BASF Process	2.7 (	"	"	)

These processes are identical to the ones selected in (2). In these processes, the amount of by-produced ammonium sulfate has drastically been decreased when compared with the Raschig process. The decrease was attained by the improvement in the production process of hydroxylamine or introducing a process which does not employ hydroxylamine. It is not economical to recover ammonium sulfate, therefore a process was developed in which gypsum is produced by decomposing the by-produced ammonium sulfate. However, in Indonesia, ammonium sulfate is utilized as a fertilizer. Therefore, the production of gypsum shall not be taken into account in this report.

(4) Others

It is difficult to carry out comparison of production cost of each process. However, concerning the above-mentioned processes such as New DSM process, BASF process, New Inventa Process and PNC process, they have been completed on the basis of a considerable extent of experiences in the production of caprolactam. Further, the amount of by-produced ammonium sulfate has been extremely decreased. Therefore, there is no significant difference in production cost among these processes.

Also, regarding the stability of operation, there is no great difference among them in view of the experiences and the actual records accomplished by the companies. However, the DSM process is a newly developed process and therefore, it is reported that the companies which employs this process, such as Nypro (U.K.), Columbia Nypro (U.S.A.) and Ube Ind. (Japan) experienced a considerable extent of operational problems.

(5) Selected Processes

The following three processes can be selected on the basis of the results of the above studies.

1. New DSM Process
2. PNC Process
3. BASF Process or New Inventa Process

Both the BASF process and the New Inventa process produce cyclohexanone by oxidizing cyclohexane and also produces hydroxylamine by the NO-Reduction process. The difference between these processes seem to exist in the reaction conditions, types and the arrangements of the equipment, etc.

3. Detailed Explanation of the Selected Processes

3-1 New DSM Process

The DSM process is a process which utilized cyclohexane as the starting raw material and, the conversion into cyclohexanone was carried out by a cyclohexane oxidation process, and further, the synthesis of a hydroxylamine was carried out by the Raschig process. After the oximation and the Beckmann rearrangement, the production of caprolactam was carried out. However, the New DSM process which involves a low extent of ammonium sulfate by-production has been developed recently and also has been industrialized by Nypro (U.K.), Columbia Nypro (U.S.A.) and Ube Ind. (Japan). The process is fully competitive, although it still has some unstable elements.

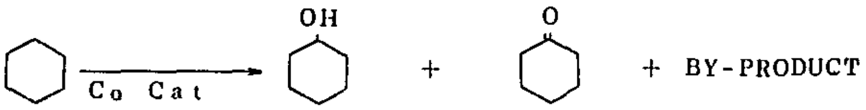
The New DSM process is analogous to the Old DSM process in that it also utilizes cyclohexane as a starting raw material and that cyclohexane is converted into cyclohexanone by oxidation. However, the outstanding characteristics of the New DSM process is that the application of the unique HPO process for the synthesis of hydroxylamine. Unlike the conventionally employed Raschig process and the NO-Reduction process, this process by-produces no ammonium sulfate at the cyclohexanone oxime producing stage. Therefore, ammonium sulfate by production, corresponding to the extent only of sulfuric acid which is used in the rearrangement of cyclohexanone oxime into caprolactam, is extremely small when compared with other processes.

The New DSM process can be divided into several steps such as oxidation of cyclohexane, oxidation of ammonia to produce hydroxylamine, production of cyclohexanone oxime from cyclohexanone and hydroxylamine, rearrangement of cyclohexanone oxime, neutralization and purification. Further, in this process, a cyclohexane oxidation process and the HPO process are incorporated. Figure X-5 shows the above process as a block diagram.

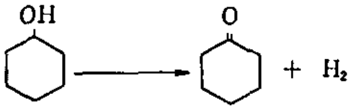
3-1-1 Chemical Reactions of New DSM Process

(1) Oxidation of Cyclohexane

At the first stage, air oxidation of cyclohexane is carried out in the liquid phase in the presence of a catalyst of cobalt-salt to produce mixed oil of cyclohexanone and cyclohexanol. A small amount of by-products is produced such as acetic acid, adipic acid, carbon monoxide, carbon dioxide, cyclohexanol, ester, etc.



This reaction is a radical oxidation reaction of cyclohexane so that the cyclohexanone and cyclohexanol are apt to be affected by a further oxidation. In order to cope with such an influence, reaction is carried out at 145 to 165°C, 8 to 10 kg/cm^2 and 4 to 6% of conversion rate per one pass. Therefore, a large amount of cyclohexane must be recycled. Cyclohexanol and cyclohexanone and unconverted cyclohexane are extracted from the reaction effluent and neutralizing the acid by caustic soda, saponifying ester, distilling cyclohexane and further, refining cyclohexanol and cyclohexanone by distillation.
Cyclohexanol is dehydrogenated to cyclohexanone.



This reaction is an endothermic reaction and it is carried out in the presence of a dehydrogenation catalyst. In this process developed by DSM, the conversion rate per one pass is 60% and the efficiency per one pass is considerably high; from 98.5% to 99%.

DSM obtained from IFP (Institut Francais du Petrole) the technology of the reactor design which utilizes boric acid as well as an air oxidation process which employs a cobalt-salt catalyst. The advantage of this process is the low consumption of cyclohexane and caustic soda and, therefore, the production cost is lower when compared with other processes. An optimum design is achieved by incorporating both the reactor design of IFP and the technology of neutralization, saponification, distillation and refining of Stamicarbon.

Cyclohexane is oxidized under the existence of boric acid and is converted into a complex compound of cyclohexanol and boric acid. The reaction temperature is 155 to 175°C and reaction pressure is 8 to 10 kg/cm^{2}G. Boric acid makes complex compounds with cyclohexanol so that a further oxidation into cyclohexanone is controlled. Therefore, the conversion rate per one pass is higher when compared with the conventional oxidation processes. In this case, it is reported that the conversion rate is approximately 10 mol% and the reaction efficiency is 90 mol%. Figure X-6 shows a block flowsheet of this process.

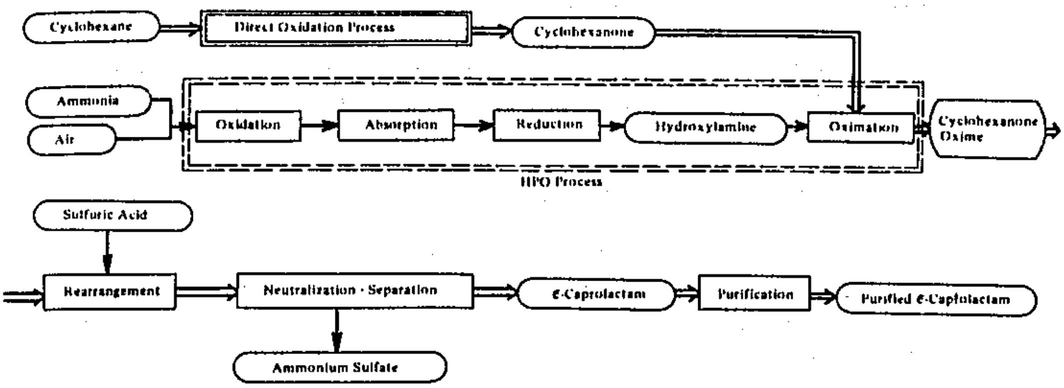
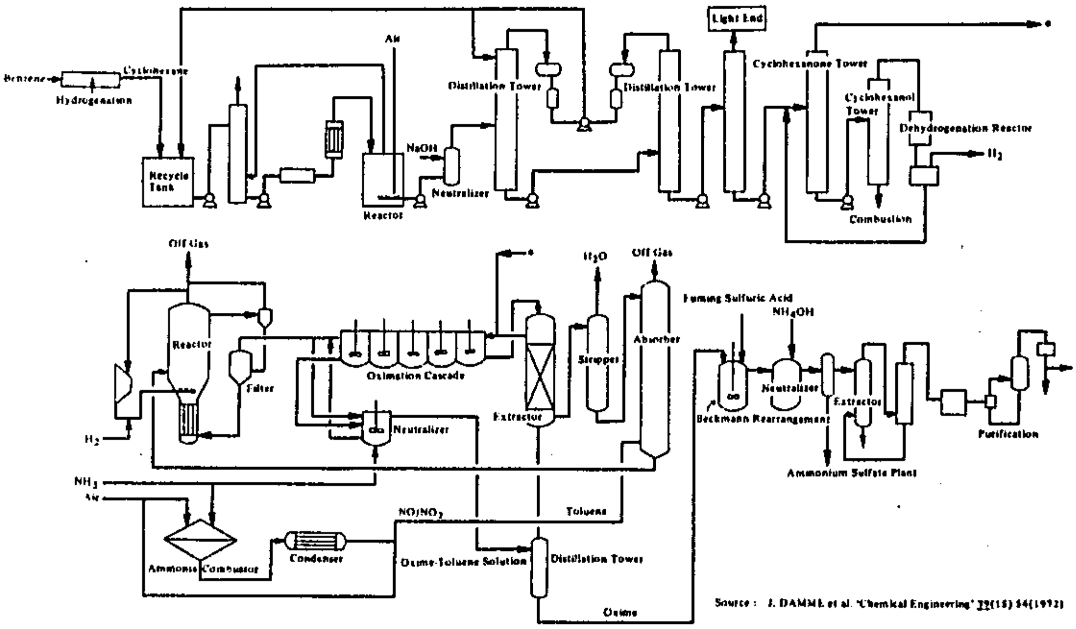


Figure X-6 Process Block Diagram of New DSM Process

(2) Formation of Hydroxylamine and Cyclohexanone Oxime

This process is the so-called HPO process which is unique to the DSM process in which the production of hydroxylamine is carried out by utilizing a noble metal catalyst and the hydrogenation of nitrate ion is carried out to produce hydroxylamine. Explanations in the following paragraphs will be made in accordance with Figure X-7.



Source : J. DAMME et al. 'Chemical Engineering' 79(18) 54(1972)

Figure X-7 Process Flow Sheet of New DSM Process

(a) Oxidation process of ammonia carries out the combustion of ammonia by air in the presence of a platinum catalyst in an oxidation reactor, thereby generating NO/NO_2 gas.

$$4NH_3 + 5O_2 \longrightarrow 4NO + 6H_2O$$

$$4NH_3 + 7O_2 \longrightarrow 4NO_2 + 6H_2O$$

$$2NO + O_2 \longrightarrow 2NO_2$$

(b) In the NO_2 absorption process, a buffer solution for phosphoric acid and ammonium phosphate (a phosphoric acid solution which has returned to the NO absorption tower after having been separated upon the completion of oximation) are employed for the absorption of NO_2 gas in the absorption tower. PH is approximately 1.8 to 2.0.

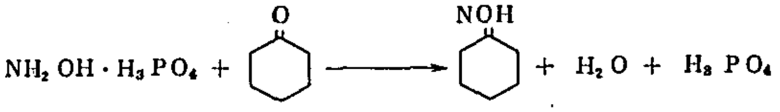
$$NH_4H_2PO_4 + HNO_3 \longrightarrow NH_4NO_3 + H_3PO_4$$

(c) The solution of ammonium nitrate and phosphoric acid which absorbed NO gas is reduced by hydrogen in a reactor in the presence of noble metal catalysts and then phosphoric acid hydroxylamine are produced.

$$NH_4NO_3 + 2H_3PO_4 + 3H_2 \longrightarrow NH_2OH \cdot H_3PO_4 + NH_4H_2PO_4 + 2H_2O$$

From the reactor, hydroxylamine phosphate and the solutions of ammonium phosphate, etc. are fed to oximation reactor, after separating catalyst.

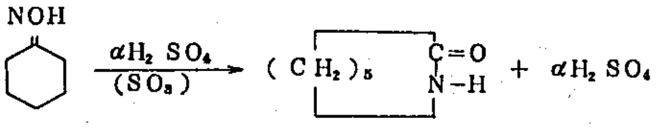
(d) In the oximation reactor, cyclohexanone oxime is produced by the following reaction.

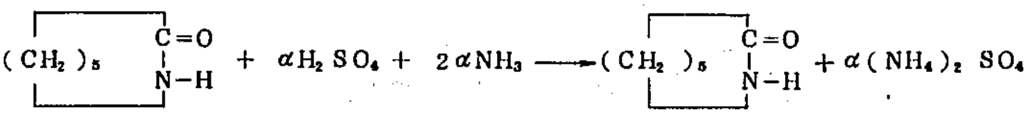


In the first stage of the reaction the reaction is carried out at PH 1 to 2 in a series of mixers and settlers and 98% of the fed cyclohexanone is oximated. In the second stage, the remaining 2% is oximated in a neutralization vessel at PH 4.5. The cyclohexanone oxime is extracted by toluene, and H_3PO_4 and $NH_4H_2PO_4$ are returned to NO_2 absorption process. After the completion of the neutralization and extraction, toluene and cyclohexanone oxime are separated by distillation and the cyclohexanone oxime is fed to rearrangement.

(e) From oxime to caprolactam

The cyclohexanone oxime which is separated from toluene at the distillation tower is converted into caprolactam by the Beckmann rearrangement in the presence of fuming sulfuric acid. This sulfuric acid plays the role of an absorbing agent against water dissolved within the oxime. The reaction heat is eliminated by cooling and the sulfuric acid contained in the reaction mixture is neutralized. The produced ammonium sulfate is removed from the caprolactam solution.





$$(CH_2)_5\begin{matrix}C=O\\|\\N-H\end{matrix} + \alpha H_2SO_4 + 2\alpha NH_3 \longrightarrow (CH_2)_5\begin{matrix}C=O\\|\\N-H\end{matrix} + \alpha(NH_4)_2SO_4$$

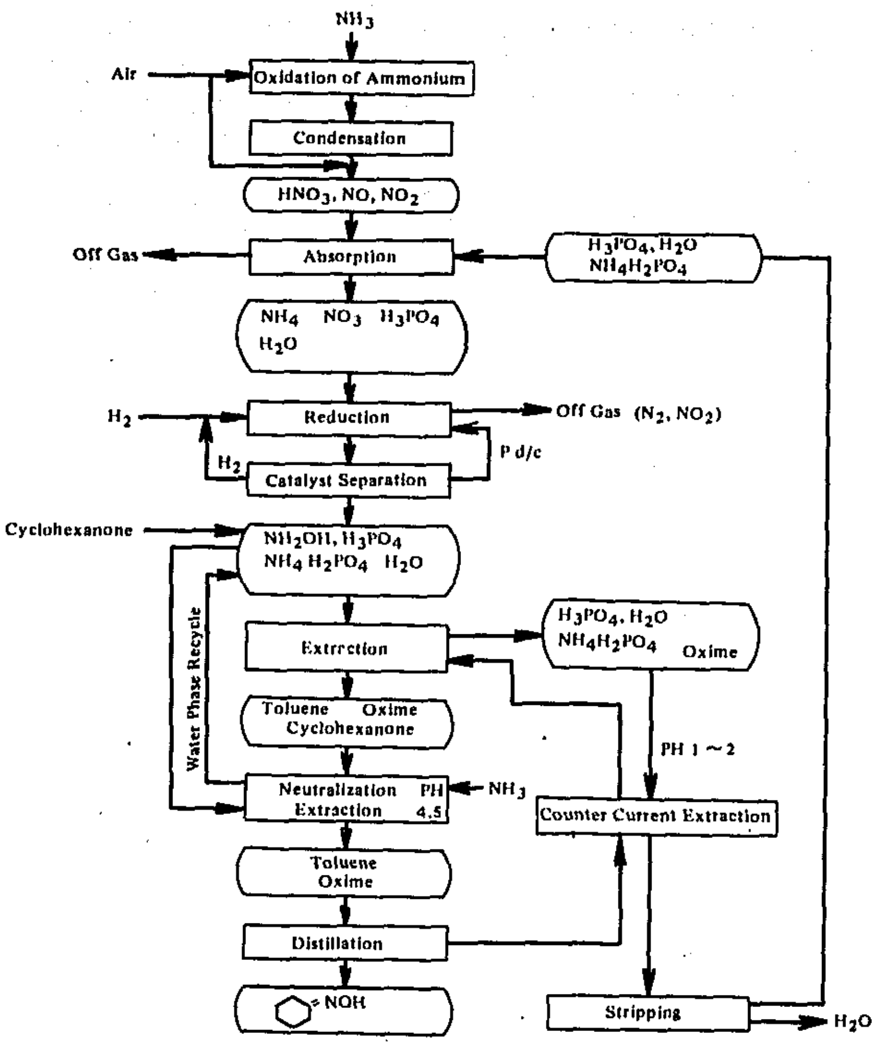
(f) The purification of caprolactam is carried out by the purification process developed by Stamicarbon and a predetermined amount of pure caprolactam is continuously produced.

3-1-2 Process Flow of New DSM Process

Figure X-8 shows the New DSM process for the production of caprolactam.

Firstly, cyclohexane is fed to the cyclohexane recycle tank of the oxidation process in which the heat exchange with the reaction off-gas is carried out. The off-gas is cooled and washed, while the water is eliminated by a separator. Thereafter, cyclohexane is heated by the heater up to the vicinity of the reaction temperature and then is fed into the reactor. The air necessary for carrying out the reaction is supplied by centrifugal compressor. Approximately 5% of cyclohexane is converted into cyclohexanol and cyclohexanone, thereby a slight amount of acid and ester are by-produced.

The liquid reaction stream from the reactor is neutralized by caustic soda and reduced to the normal pressure after passing through the intermediate storage tank. The caustic soda solution in the neutralization vessel is fed to the steam-distillator and the dissolved cyclohexanol and cyclohexanone is recovered therein. The cooled oxidation products coming out from the cyclohexane distillation tower is fed to the saponifier in which the cyclohexanol ester is hydrolized by caustic soda. A slight amount of sodium salt is washed and eliminated in the extraction tower. Further, the remaining cyclohexane is eliminated by distillation and then recycled to the reaction process.



Source : J. DAMME et al. "Chemical Engineering" 79(15) 54 (1972)

Figure X-8 Process Flow of New DSM Process

The mixture of the crude cyclohexanol and cyclohexanone are fed into the cyclohexanone tower after the elimination of the light-fractions. In the tower, it is separated into cyclohexanol and cyclohexanone. The cyclohexanone is fed to the oximation reactor. Cyclohexanol on the other hand is fed into the cyclohexanol tower in which the elimination of the high-boiling point substances is conducted and then fed to the catalytic reactor and converted to cyclohexanone. Dehydrogenation reactor effluent is condensed and then separated from hydrogen and thereafter return-fed to the cyclohexanone tower in which they are distilled and cyclohexanone is produced.

In the following paragraphs, an explanation is made concerning the processes covering a range from the hydroxylamine up to the oxime.

Firstly, ammonia and air are combusted by means of a blower while passing through the NH_3 oxidizer and after the water content is removed by means of a condenser, the NO_2 gas is led to the absorption tower. In the absorption tower, the buffer solution (a mixture of phosphoric acid and ammonium phosphate) which is recycled from the dehydrating tower absorb NO_2 to produce ammonium nitrate/phosphoric acid solution. This solution is then fed to the reactor in which the reduction by hydrogen is carried out to produce hydroxylamine phosphate and ammonium phosphate. The solution coming out from the reactor is fed to the oxime reactor after the separation of the catalyst. Cyclohexanone, hydroxylamine phosphate, ammonium phosphate and toluene are fed into oximation reactor and the oximation reaction proceeds under the counter current conditions. The cyclohexanone oxime is extracted into toluene. No clarification is made concerning such details as the reaction temperature, pressure, etc. The solution is neutralized by ammonia at PH 4.5 in a neutralization vessel. Approximately 3 vol.% of the side stream from the main stream and NH_3 are fed to neutralizer. The oxime extracted into toluene separated into toluene and cyclohexanone oxime by distillation and the separated cyclohexanone oxime is fed to the subsequent Beckmann rearrangement stage. On the other hand, the phosphoric compounds which is the recycling buffer solution coming out from the oxime reactor is fed into the extraction tower in which the remaining cyclohexanone oxime is extracted. After the extraction, the compound is led to the dehydrating tower and then further fed to the NO_2 absorption tower and recycled into reduction reactor. Production of caprolactam from cyclohexanone oxime is carried out by the following process stages.

Fuming sulfuric acid is added to cyclohexanone oxime to produce caprolactam by Beckmann rearrangement of cyclohexanone oxime. The fuming sulfuric acid is playing the role of an absorbing agent against the water dissolved into the oxime. The reaction heat is eliminated by cooling and the rearranged reaction compounds are neutralized by ammonia in the subsequent neutralization vessel, thereby by-producing ammonium sulfate. The separation of ammonium sulfate from caprolactam solution is carried out by the specific gravity difference of the solution in the separator.

The crude caprolactam is then purified continuously by the process developed by Stamicarbon and pure caprolactam is produced. The crude caprolactam is extracted by benzene and further re-extracted by water. After passing through a series of physical-chemical purification stages, the whole process is completed. The water solution of caprolactam is concentrated through a series of evaporator and finally made into either flaked caprolactam or water containing caprolactam.

Table X-6 shows the utilities unit consumption and the raw materials unit consumption for the New DSM process.

3-2 PNC Process.

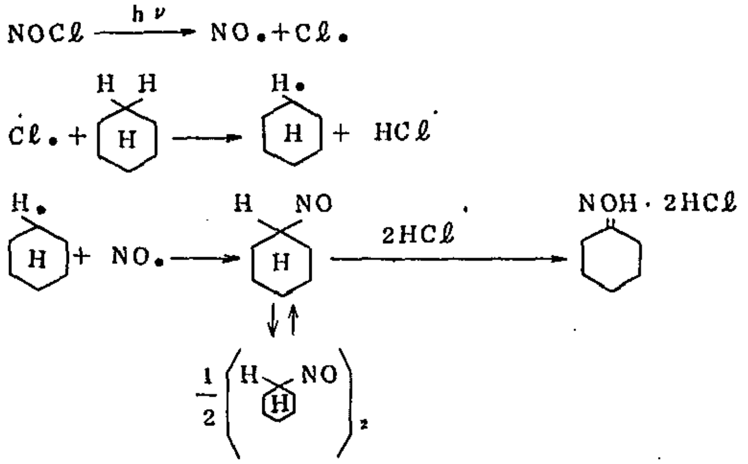
The PNC process was developed by Toray as a caprolactam production process by which it is possible to produce cyclohexanone oxime by means of the photonitrosation of cyclohexane. Toray has so far expanded the facilities as follows. The production was 40 t/d in 1963, 140 t/d in 1966, 255 t/d in 1968 and 380 t/d in 1971. Thus, Toray has entirely switched to the PNC process from the conventional facilities in which the phenol process as well as the so-called cyclohexanone intermediate production process. At present the expansion of the facilities has been made up to the present production capacity of 150,000 t/y.

3-2-1 Chemical Reactions of PNC Process

The main portion of this process consists of the cyclohexanone oxime synthesis process by means of photonitrosation reactions.

The nitrosyl chloride is dissociated by light into chlorine radical and nitrogen oxide radical. A high output lamp (40 to 60 KW of wave-length 400 to 600 mμ) is employed as the light source. The chlorine radical abstract hydrogen from cyclohexane, thereby separating cyclohexane into hydrogen chloride and cyclohexyl radical. The cyclohexyl radical will combine itself with nitrogen oxide, thereby producing nitrosyl cyclohexane. However, nitrosyl cyclohexane is not stable under the photo-reaction conditions so that it changes into bis-nitrosyl cyclohexane by dimerization, or into cyclohexanone oxime hydrochloride in the presence of hydrogen chloride.

Cyclohexanone oxime hydrochloride is an oily solution which does not dissolve in cyclohexane, therefore, it is precipitated inside the reaction vessel and separated.



The following points can be enumerated as the features of the PNC process.

(1) While a comparatively high-cost hydroxylamine is employed for the introduction of nitrogen in the cyclohexanone oxime when employing the air-oxidation process, the PNC process can obtain cyclohexanone oxime by employing a comparatively low cost nitrosyl chloride. A number of routes are available for the synthesis of nitrosyl chloride. The following reaction processes are suitable for commercial production of nitrosyl chloride.

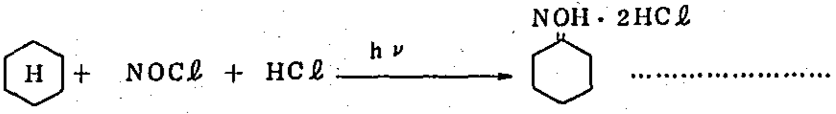
$$2NH_3 + 3O_2 \longrightarrow N_2O_3 + 3H_2O \quad \cdots\cdots\cdots \quad (1)$$

$$2H_2SO_4 + N_2O_3 \longrightarrow 2HNOSO_4 + H_2O \quad \cdots\cdots \quad (2)$$

$$HNOSO_4 + HC\ell \longrightarrow H_2SO_4 + NOCI \quad \cdots\cdots \quad (3)$$

The H_2O evolved in the reaction of the above formula(2) is separated under reduced pressure and heating. Therefore, it is possible to carry out the reutilization of sulfuric acid.

(2) A direct oxime production by means of photo-chemical reaction can be undertaken from cyclohexane without going through the cyclohexanone stage. The yield in this process is higher than that of the air-oxidation process because of the unnecessity for the cyclohexane oxidation process.



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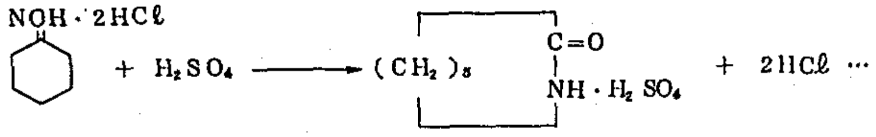


(5)

(3) The reaction of the above chemical formula (5) is approximately the same as the Beckmann rearrangement of cyclohexane oxidation process. However, because of the generation of hydrogen chloride gas, the extent of the heat of reaction is approximately 1/2 of that of conventional reaction, thereby making the removal of heat easy. The evolved hydrogen chloride gas is recovered in the form of gas. The product of this reaction is the mixture of caprolactam and sulfuric acid and sent to neutralization and purification processes, which are virtually the same as the process which goes through the cyclohexanone process.

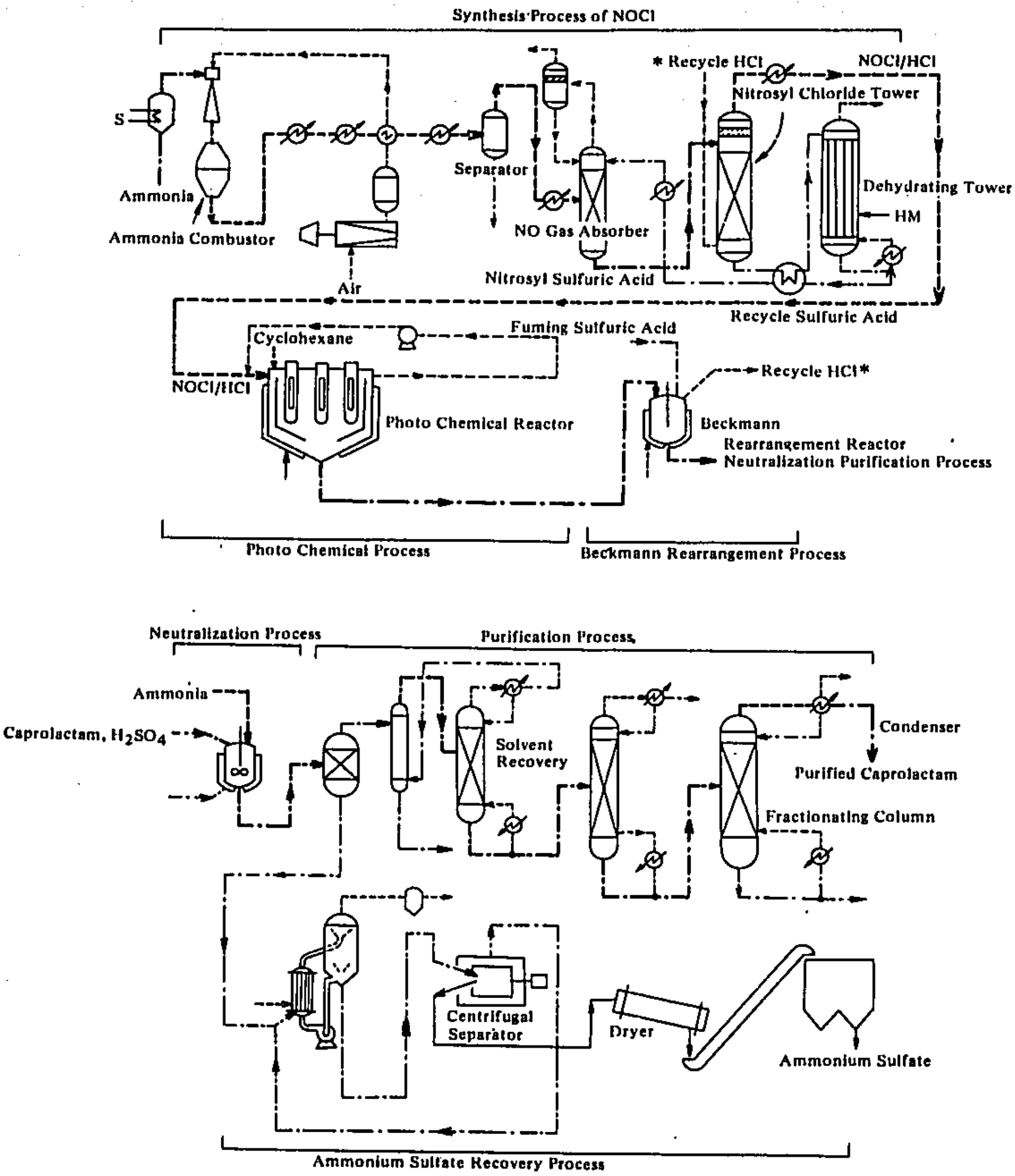
In the PNC process, no ammonium sulfate is by-produced except at the time of neutralization step after the Beckmann rearrangement, thereby making this process advantageous in view of the low extent of by-production of ammonium sulfate which amounts only to 1.7 kg for a kg of caprolactam.

3-2-2 Process Flow of PNC Process

The outline of the process is shown in Figure X-9.

The reactions from ammonia up to nitrosyl chloride are as shown in the formulae (1) through (3). Ammonia is almost quantitatively oxidized by platinum-net catalyst. The nitrosyl sulfuric acid producing reactions represented by the formula (2) also progresses almost quantitatively. In this reaction, such inert gas as nitrogen gas, etc., are separated which have been contained in the air used to oxidize ammonia.

The reaction represented by the chemical formula (3) is carried out by a counter-current contact with the catalyst inside the filling tower. In this tower, the waste acid is dehydrated in the dehydration tower and then is recycled to the nitrosyl sulfuric acid production process. The obtained mixture gas of nitrosyl chloride and hydrogen chloride is fed to the photo-chemical reaction by means of a blower.



Source : Process Handbook
(The Japan Petrochemical Institute)

Figure X-9 Process Flow Sheet of PNC Process

Buffer boards are provided inside the photo-reaction vessel which also undertakes the cooling. The light source is submerged into the solution together with the cooling outer tubes and the supplied NOCl/HCl mixed gas is blown from outside the buffer boards and then finally supplied through nozzles. In order to control the sub-reaction, the reaction temperature is maintained at 10 to 20°C by the function of the jacket as well as by means of feeding a coolant inside the buffer boards. The reacted substance is the salt of 2-mol of hydrochloric acid and cyclohexanone oxime. As the salt is not soluble in cyclohexanone, it is sedimentated on the bottom of the reaction vessel in the form of an oily liquid and continuously withdrawn.

Glass linings or titanium is employed as the structural material to form the photo-reaction vessel. The capacity per unit of the photo-chemical reactor can be controlled by changing the number of the lamps to be inserted. However, the optimum level is 10 to 50 t/d in view of economic considerations.

The 2-mol hydrochloric acid salt of cyclohexanone oxime which is generated by the photo-chemical reaction is continuously fed into the Beckmann rearrangement reaction vessel in which it is changed into caprolactam sulfuric acid solution.

Further, in the case of the PNC process, the discharging of hydrogen chloride is undertaken at the Beckmann rearrangement stage so that the generation of reaction heat is much lower than the case of employing cyclohexanone oxime alone, thereby making the temperature control easy. The reaction progresses quantitatively, thereby hydrogen chloride is recovered almost completely.

The rearrangement reaction solution is neutralized by ammonia and separates into the caprolactam layer and the ammonium sulfate layer, the crude caprolactam aqueous solution can be obtained by separating this solution. Ammonium sulfate is also obtained in the form of a aqueous solution and recovered in the form of crystallized ammonium sulfate by condensation and crystallization.

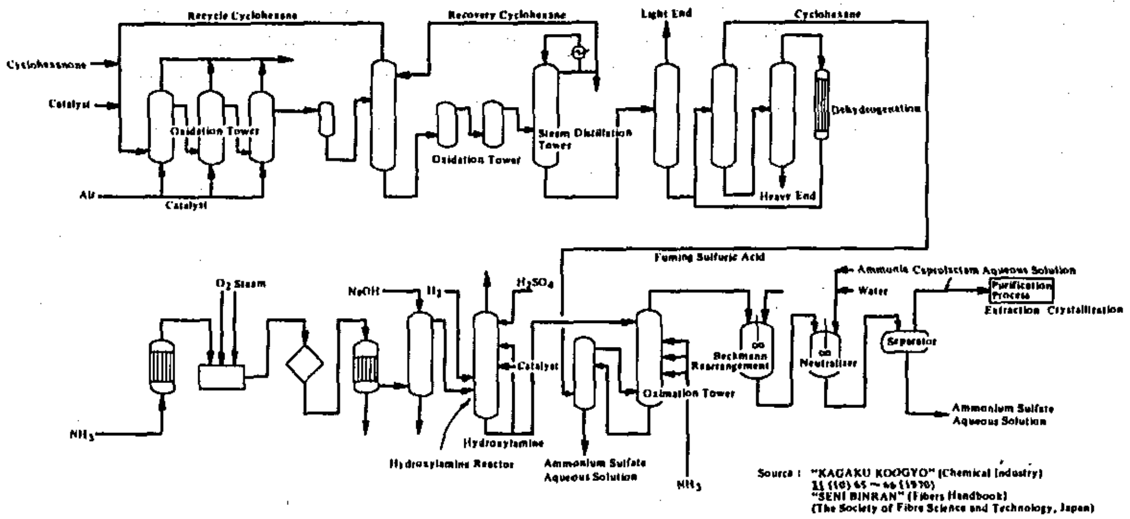
Table X-9 shows the raw materials unit consumption and utilities unit consumption in the case of the PNC process.

3-3 BASF Process and New Inventa Process

Both the BASF process and the New Inventa process employ the cyclohexane oxidation process and NO-Reduction process, thereby presenting no basic difference between the two. The difference between the two processes exist in the types of catalyst to be used and the reaction conditions as well as in the employed equipment so that the BASF process will be taken as an example for proceeding with the process description.

The BASF process is a well-known process which employs the oxidation of cyclohexane. The production of hydroxylamine is carried out by means of the NO-Reduction process.

3-3-1 Process Description (Refer to the Flowsheet in Figure X-10)



Source : "KAGAKU KOOGYO" (Chemical Industry) 21 (10) 65 ~ 66 (1970)
"SENI BINRAN" (Fibers Handbook)
(The Society of Fibre Science and Technology, Japan)

Figure X-10 Process Flow Sheet of BASF Process

(1) Production of Cyclohexanone

Basically, the employed process is the cyclohexane oxidation process. Therefore, there is no basic difference from the DSM process, etc. In the following paragraphs, the points of difference from the DSM process will be described.
Cyclohexane is oxidized by air to produce cyclohexanone and cyclohexanol. As to the catalyst, it is assumed that cobalt naphthenate is employed. (It is reported that BASF is also employing a process in which boron or boric acid is used as catalyst.) It is also assumed that reaction conditions are 140°C at 25 atm.

The BASF process carries out washing of three reactors separately by the downstream to remove the acid after the completion of the air oxidation. The KA oil from the reactor is fed to decomposition tower and then to recovery tower. In the recovery tower, cyclohexane is separated and returned to the reaction cycle. The mixed oil discharged from the recovery tower is fed to the saponifier and the ester is saponified therein. Cyclohexane is recovered by steam distillation and cyclohexanone and cyclohexanol are separated after removing the low-boiling-point substances. The cyclohexanol is further converted into cyclohexanone in the dehydration tower after the removal of impure substances. Then, it is fed to the cyclohexanone distillation tower and added to the main flow of cyclohexane.

(2) Production of Hydroxylamine

This is the NO-Reduction process in which the amount of by-produced ammonium sulfate is reduced to 1/2 of that in the conventional Raschig process. In other words, ammonia is oxidized to NO by pure oxygen under the existence of steam and then NO is reduced by hydrogen in the presence of platinum catalyst in sulfuric acid (a high quality of 96 to 98% concentration is required) to produce hydroxylamine sulfate. The main reactions are as follows:

$$4NH_3 + 5O_2 \longrightarrow 4NO + 6H_2O$$

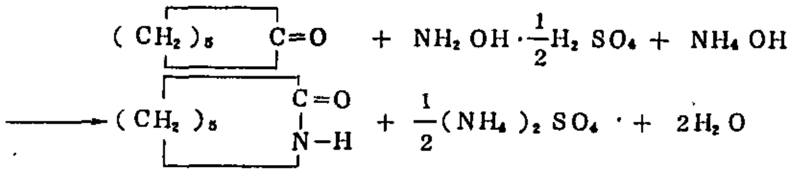
$$NO + \frac{3}{2}H_2 + \frac{1}{2}H_2SO_4 \longrightarrow NH_2OH \cdot \frac{1}{2}H_2SO_4$$

In this process, it is necessary to employ high purity oxygen and sulfuric acid. In the case of the Raschig process, the oxidation of ammonia is carried out by air to obtain N_2O_3, however, in this process, the oxidation is carried out by using pure oxygen. The reason for using pure oxygen is that, if NO alone is intended to be obtained, it would be necessary to supply oxygen stoichiometrically so that the completion of the reactions would not be achieved unless pure oxygen is used. However, because of the fact that the use of oxygen and ammonia alone bring the reaction into the explosion range, it is necessary to provide the coexistence of steam at a certain rate. Platinum nets are used as the catalyst.

The reactions are carried out at more than 70°C and the reaction accompany a large extent of heat generation. Nitric acid can be recovered as a by-product. The reduction of NO is carried out in sulfuric acid solution using platinum as the catalyst. For instance, mixed gas of 60% hydrogen and 40% NO are fed to tower-type reactor to obtain hydroxylamine sulfate. The reaction temperature is approximately on a level of 75° C and it is reported that the yield is 87% to 90%.

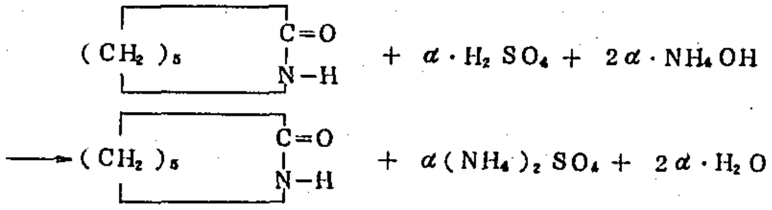
(3) From Oxime to Caprolactam

Cyclohexanone is oximated by hydroxylamine sulfate in the oxime reaction vessel. At this time, sulfuric acid reacts with ammonia, thereby producing ammonium sulfate.

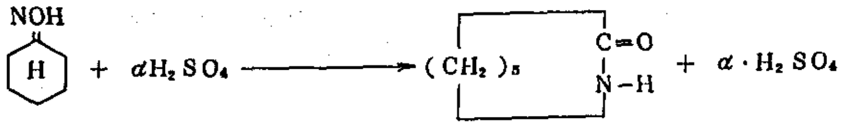


$$(CH_2)_5C{=}O + NH_2OH \cdot \frac{1}{2}H_2SO_4 + NH_4OH \longrightarrow (CH_2)_5\begin{matrix}C{=}O\\|\\N{-}H\end{matrix} + \frac{1}{2}(NH_4)_2SO_4 + 2H_2O$$

Therefore, in the case of the BASF process, the by-produced amount of ammonium sulfate is 1/2-mol as against 1-mol of cyclohexanone oxime at this process stage, which is 1/2 of the by-production compared with the case of the conventional Raschig process. This by-produced ammonium sulfate is fed to the ammonium sulfate purification process as an aqueous solution. The cyclohexanone oxime is sent to a reactor to undergo the Beckmann rearrangement in the presence of fuming sulfuric acid, thereby producing ε-caprolactam.



$$(CH_2)_5\begin{matrix}C{=}O\\|\\N{-}H\end{matrix} + \alpha \cdot H_2SO_4 + 2\alpha \cdot NH_4OH \longrightarrow (CH_2)_5\begin{matrix}C{=}O\\|\\N{-}H\end{matrix} + \alpha(NH_4)_2SO_4 + 2\alpha \cdot H_2O$$



(4) Caprolactam Purification

The purification of caprolactam is carried out by extraction, crystallization, etc. so that there is no specific difference from the other processes.

The raw materials unit consumption of the BASF process are shown in Table X-7.

3-4 Comparison of the Unit Consumption in Various Processes

The unit consumption varies from process to process. As it is highly difficult to carry out economic evaluation for each one of the various processes, a typical example will be formulated in the following paragraphs and the subsequent economic calculations will be made on the basis of this example.

Table X-13 shows the results of comparative studies of the unit consumption of the selected processes. Unit consumption for cyclohexane for the PNC process is 0. 908 which is evidently smaller than the others. This is due to the fact that no cyclohexane oxidation process is necessary for the PNC process.

Table X-13 Comparison of Unit Consumption in Various Caprolactam Processes

		New DSM 1)	PNC	BASF	New Inventa	Standard Case
Raw Materials						
Cyclohexane	kg	(1.06) 2)	0.908	1.150	1.01	1.0
Ammonia	kg	0.805	0.649	0.953	0.96	0.688
Sulfuric Acid	kg	1.368	1.279	1.97	2.05	1.282
Hydrogen	m^3	(0.914) 3)		0.467	0.47	
Oxygen	m^3			0.559	0.38	
Catalysts & Chemicals	US¢					5
By Product						
Ammonium Sulfate	kg	-1.8	-1.7	-2.7	-2.6	-1.7
Utilities						
Electiricity	kwh	(0.45) 4)	4.05		0.9	
Steam	kg	(12.65) 4)			12	
Cooling Water	m^3	(1.62) 4)			0.14	
Fuel	10^3 kcal	(0.47) 4)			3	1.38 kg

Notes: 1) As no data are available concerning New DSM from cyclohexane, the above data are estimated from the data of Old DSM and New DSM from benzene or cyclohexane.

2) Data for Old DSM

3) When caprolactam is produced from benzene hydrogen consumption is reported to be 1.814 m^3 and 0.9 m^3 is consumed to hydrogenation of benzene. Hydrogen consumption from cyclohexane is estimated at 0.914 m^3.

4) Data when benzene is used as raw materials.

Naturally, the unit consumption for sulfuric acid and ammonia is smaller when the by-produced amount of ammonium sulfate is lower. The hydrogen consumption by the New DSM process displays approximately twice the level of that required by the BASF process or by the New Inventa process. However, this is due to the fact that in the New DSM process, a large amount of hydrogen is required in the hydroxylamine production process. In the cases of the BASF process and New Inventa process, oxygen is utilized at the hydroxylamine production stage. In addition to this, catalysts and other chemicals are necessary.

Typical unit consumption data by summarizing these factors are also shown in Table X-13. Concerning ammonium sulfate, the value of 1.7 which is the lowest extent of the by-production rate has been employed in this Table. Also, regarding the utilities, reference was made to the New DSM and New Inventa processes and are stipulated in terms of the fuel unit consumption after incorporating the data concerning electrical power, steam, etc.

Due to the fact that the PNC process consumes approximately 4 KWH of electrical power for carrying out the photo-chemical reactions, the 1.38 kg of fuel shown in terms of the standard unit consumption will evidently be insufficient. Also the cost for replacing lamps will be inevitable. However, in view of the low extent of the cyclohexane unit consumption and the absence of the cyclohexane oxidation process and, further, the production of simply producible nitrosyl chloride is carried out instead of the comparatively complicated hydroxylamine production, it was assumed that such a shortcoming will be cancelled in view of the total production cost consideration.

4. Construction and Operation of a Plant

4-1 Construction Schedule

Figure X-11 shows a tentative schedule compiled on a target of the completion of the plant construction by mid-1977 and the commencement of the test operation immediately after the construction completion.

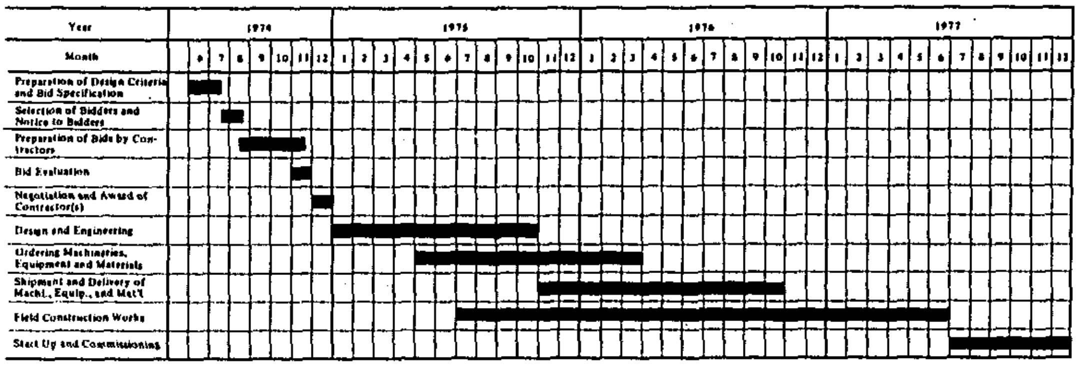


Figure X-11 Tentative Schedule of 60,000 t/y Caprolactam Plant Construction

4-2 Estimate on the Construction Cost

Detailed explanations have already been made in a clause pertaining to the Survey Method regarding the scope and the contents of the estimate to be made on the construction cost.

Tables X-14 and X-15 show the details of the investment amount required in the case of constructing a caprolactam plant in Indonesia by mid-1977. The total investment amount will be US$124 million in the case of 60,000 t/y caprolactam production and US$95 million in the case of 40,000 t/y production. The estimate was firstly compiled on an assumption of constructing the plant in Indonesia in the year 1971, to which the escalation and contingency were added in order to obtain the total plant cost required in the year 1976. The transportation includes the ocean freight and marine insurance premium from Japan to Indonesia and the total plant cost includes the interest during construction, royalty, engineering fee, technical expenses, land cost and working capital in order to obtain the total investment amount at the time of 1977. Detailed explanations have already been made concerning the computing method of the amount for each one of these items in the clause pertaining to the Survey Method.

Table X-14 Estimated Capital Requirement (1977) Case 1

Case 1 Caprolactam 60,000 t/y (Base Year 1971)

	Rupiah (10^6)			U.S.$ Equiv. (10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchange Portion	Local Currency Portion	Total
1. Equipment						
Plant	7,698	-	7,698	18,549	-	18,549
Utilities Plant & Facilities	4,001	-	4,001	9,642	-	9,642
Auxiliary & Off-Site	1,541	-	1,541	3,714	-	3,714
Sub-Total	13,240	-	13,240	31,905	-	31,905
2. Construction Costs						
Installation	1,273	748	2,021	3,067	1,803	4,870
Civil Works	1,258	1,537	2,795	3,030	3,702	6,732
Office Bldg. & Housing Colony	272	759	1,031	656	1,828	2,484
Sub-Total	2,803	3,044	5,847	6,753	7,333	14,086
3. Transportation	3,575	-	3,575	8,614	-	8,614
4. Plant Cost	19,618	3,044	22,662	47,272	7,333	54,605
5. Escalation & Contingency	13,384	2,076	15,460	32,250	5,003	37,253
6. Total Plant Costs	33,002	5,120	38,122	79,522	12,336	91,858
(1. Process Plant)	18,587	1,259	19,846	44,788	3,033	47,821
(2. Auxiliary & Off-Site)	13,957	2,585	16,542	33,631	6,228	39,859
(3. Buildings)	458	1,276	1,734	1,103	3,075	4,178
7. Royalty, Eng. Fee, Technical Exp.	2,789	-	2,789	6,721	-	6,721
8. Pre-Operating Expenses	-	630	630	-	1,518	1,518
9. Land	-	725	725	-	1,747	1,747
10. Interest During Construction	2,774	-	2,774	6,684	-	6,684
11. Working Capital	-	6,375	6,375	-	15,361	15,361
12. Total Capital Requirement	38,565	12,850	51,415	92,927	30,962	123,889

Table X-15 **Estimated Capital Requirement (1977) Case 2**

Case 2 Caprolactam 40,000 t/y (Base Year 1971)

	Rupiah (10^6)			U.S.$ Equiv. (10^3)		
	Foreign Exchange Portion	Local Currency Portion	Total	Foreign Exchange Portion	Local Currency Portion	Total
1. Equipment						
Plant	5,812	-	5,812	14,003	-	14,003
Utilities Plant & Facilities	3,194	-	3,194	7,697	-	7,697
Auxiliary & Off-Site	1,160	-	1,160	2,794	-	2,794
Sub-Total	10,166	-	10,166	24,494	-	24,494
2. Construction Costs						
Installation	1,003	578	1,581	2,417	1,392	3,809
Civil Works	1,035	1,266	2,301	2,494	3,050	5,544
Office Bldg. & Housing Colony	219	705	924	528	1,700	2,228
Sub-Total	2,257	2,549	4,806	5,439	6,142	11,581
3. Transportation	2,745	-	2,745	6,614	-	6,614
4. Plant Cost	15,168	2,549	17,717	36,547	6,142	42,689
5. Escalation & Contingency	10,347	1,738	12,085	24,933	4,189	29,122
6. Total Plant Costs	25,515	4,287	29,802	61,480	10,331	71,811
(1. Process Plant)	14,102	972	15,074	33,980	2,342	36,322
(2. Auxiliary & Off-Site)	11,044	2,129	13,173	26,611	5,131	31,742
(3. Buildings)	369	1,186	1,555	889	2,858	3,747
7. Royalty, Eng. Fee, Technical Exp.	2,099	-	2,099	5,056	-	5,056
8. Pre-Operating Expenses	-	420	420	-	1,013	1,013
9. Land	-	591	591	-	1,425	1,425
10. Interest During Construction	2,160	-	2,160	5,204	-	5,204
11. Working Capital	-	4,501	4,501	-	10,846	10,846
12. Total Capital Requirement	29,774	9,799	39,573	71,740	23,615	95,355

Regarding the estimate on the construction cost, it is possible that a certain extent of discrepancy will take place depending upon the type of the process to be employed. However, concerning this point, an average value was calculated to be applied for the studies to be made in this report. Concerning the extent of the influence to be exerted on the economic viability of the project by the variation in the construction cost, sensitivity analysis have been made in the clause pertaining to the Economic Evaluation.

4-3 Number of Personnel

The direct and indirect number of required people for carrying out the plant operation is as shown in Table X-16. The total required number of people is 400 in the case of 60,000 t/y operation and 349 in the case of 40,000 t/y operation.

Table X-16 Required Number of Workers for Caprolactam Plant (Direct, Indirect and Total)

(person)

	60,000 t/y	40,000 t/y
Works Manager	1	1
Manager	2	2
Unit Superintendent	7	7
Section Superintendent	32	27
Foreman	40	35
Operator	236	205
Labourer	82	72
Total	400	349

4-4 Operation Training

For carrying out the plant operation, one unit superintendent, 14 section superintendents and 19 foremen, totaling 34 persons per year approximately will be necessary. These personnel should consist of experienced expatriate engineers.

4-5 Plant Layout

Figure X-12 shows the caprolactam plant layout for 60,000 t/y operation. The required total area except the housing colony will be 143,100 m^2.

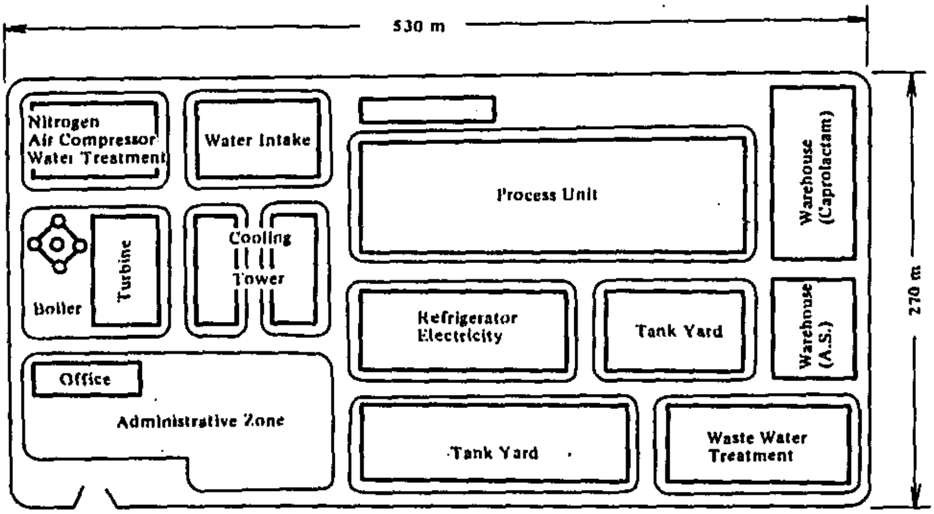


Figure X-12 Plant Layout of 60,000 t/y Caprolactam Production

5. Economic Evaluation

5-1 Production and Operational Rate

An assumption is made here that the commencement of test operation be made in July 1977 and the production amount for the first year operation including the 3-month test operation period shall be 1/4 of the whole production capacity. From the second year onward, it is assumed that the full-capacity production operation will be carried out.

5-2 The Amounts of Shipment to the Domestic Market and the Export Market

The domestic demand amount has already been shown.

The shipment amount for domestic market will be obtained on the basis of the correlation between the domestic demand extent and production amount. The balance between shipment amount for domestic market and production amount is assumed to be allocated for exportation. The summary of the results of such calculations are shown in Tables X-17 and X-18. Concerning the year 1977, the domestic demand amounts to 26,000 t/y, however, due to the fact that the commencement of the operation is targeted for July 1977, the supply of the products will be possible only for covering the demand generated after July 1977. Therefore, the domestic demand extent is estimated as being 13,000 t/y. Regarding the demand from 1982 onward, the demand growth extent from 1980 to 1981 was directly applied for estimation.

Table X-17 Production, Domestic shipment and Export shipment of Caprolactam (60,000 t/y)

(10^3 t/y)

	Production	Domestic Demand	Consumption		
			Domestic	Export	Total
1977	15	26	13	2	15
78	60	33	33	27	60
79	60	41	41	19	60
80	60	49	49	11	60
81	60	51	51	9	60
82	60	(53)	(53)	7	60
83	60	(55)	(55)	5	60
84	60	(57)	(57)	3	60
85	60	(59)	(59)	1	60
86	60	(61)	(60)	0	60

Table X-18 Production, Domestic Shipment and Export Shipment of Caprolactam (40,000 t/y)

(10^3 t/y)

	Production	Domestic Demand	Consumption		
			Domestic	Export	Total
1977	10	26	10	0	10
78	40	33	33	7	40
79	40	41	40	0	40
80	40	49	40	0	40
81	40	51	40	0	40
82	40	(53)	40	0	40
83	40	(55)	40	0	40
84	40	(57)	40	0	40
85	40	(59)	40	0	40
86	40	(61)	40	0	40

The comparison between the caprolactam 60,000 t/y and 40,000 t/y production cases reveals that the former involves a higher rate of exportation, i.e., 45% in 1978, 32% in 1979 and 18% in 1980, however, in the case of the latter, the total production will be entirely allocated for domestic consumption from 1979 onward, thereby making it necessary to either carry out importation or to construct a new plant in order to cover the demand generated the years thereafter.

5-3 Production Cost Calculation

By utilizing the standard unit consumption shown in Table X-13, the 1977 prices summarized in the clause pertaining to the Price Forecast, and the investment amount shown in Tables X-14 and 15, calculations of the average production cost was undertaken. The production costs in the case of 60,000 t/y production and 40,000 t/y production are shown in Tables X-19 and X-20. The production cost after interest in the case of 60,000 t/y operation was US¢88/kg, while in the case of 40,000 t/y, US¢93.6/kg, thereby showing a discrepancy by US¢5.6/kg.

Table X-19 Production Cost of Caprolactam (60,000 t/y)

Table X-19-1 Estimated Production Cost of Caprolactam

	Unit Consumption (kg/kg)	Price (US¢/kg)	Unit Cost (US¢/kg)
Variable Costs			
Raw Materials			
Cyclohexane	1.0	30.5	30.50
Sulfuric Acid	1.282	4.5	5.76
Ammonia	0.688	15.0	10.32
Catalysts and Chemicals			5.00
Ammonium Sulfate	1.7	-7.3	-12.41
Total Raw Material Costs			39.17
Utility			
Fuel	1.38	8.3	11.45
Total Utility Costs			11.45
Total Variable Costs			50.63
Wages			1.22
Depreciation			
Battery Limits			7.97
Off-Site			4.42
Building			0.23
Royalty, Eng. Fee			2.24
Pre-Operating Expenses			0.50
Int. During Construction			2.22
Total Depreciation			17.60
Repairs and Insurance			6.12
Taxes			0
Plant Overhead Costs			1.22
Total Fixed Costs			26.17
Running Royalty			0
Factory Costs			76.80
Selling Expenses			1.40
General Administrative Exp.			2.30
Total Product. Cost Before Int.			80.50
Interest			
Int. on Total Inv. Costs			4.38
Int. on Working Capital			3.07
Total Interest			7.46
Total Product. Cost Including Int.			87.97

Working Capital (10^3 US$)	15,361
Total Investment Cost *(10^3 US$)	123,889

* Including Land Price & Working Capital

Table X-19-2 Production Cost

(10³ US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production (10^3 ton)	15	60	60	60	60	60	60	60	60	60
Variable Costs										
Raw Materials										
Cyclohexane	4,575	19,581	20,951	22,418	23,987	25,666	27,463	29,385	31,442	33,643
Sulfuric Acid	865	3,703	3,962	4,240	4,537	4,854	5,194	5,558	5,947	6,363
Ammonia	1,548	6,625	7,089	7,585	8,116	8,684	9,292	9,942	10,639	11,383
Catalysts and Chemicals	750	3,210	3,434	3,675	3,932	4,207	4,502	4,817	5,154	5,515
Ammonium Sulfate	-1,861	-7,967	-8,524	-9,121	-9,760	-10,443	-11,174	-11,956	-12,793	-13,689
Total Raw Material Costs	5,876	25,152	26,913	28,797	30,813	32,970	35,278	37,747	40,390	43,217
Utility										
Fuel	1,718	7,353	7,868	8,418	9,008	9,638	10,313	11,035	11,808	12,634
Total Utility Costs	1,718	7,353	7,868	8,418	9,008	9,638	10,313	11,035	11,808	12,634
Total Variable Costs	7,594	32,506	34,781	37,216	39,821	42,609	45,591	48,783	52,198	55,852
Wages	733	784	839	897	960	1,028	1,100	1,177	1,259	1,347
Depreciation										
Battery Limits	4,782	4,782	4,782	4,782	4,782	4,782	4,782	4,782	4,782	4,782
Off-Site	2,657	2,657	2,657	2,657	2,657	2,657	2,657	2,657	2,657	2,657
Building	139	139	139	139	139	139	139	139	139	139
Royalty, Eng. Fee	1,344	1,344	1,344	1,344	1,344	0	0	0	0	0
Pre-Operating Expenses	303	303	303	303	303	0	0	0	0	0
Int. During Construction	1,336	1,336	1,336	1,336	1,336	0	0	0	0	0
Total Depreciation	10,563	10,563	10,563	10,563	10,563	7,578	7,578	7,578	7,578	7,578
Repairs and Insurance	3,674	3,674	3,674	3,674	3,674	3,674	3,674	3,674	3,674	3,674
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	733	784	839	897	960	1,028	1,100	1,177	1,259	1,347
Total Fixed Costs	15,703	15,806	15,916	16,033	16,159	13,309	13,453	13,607	13,771	13,948
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	23,298	48,312	50,697	53,250	55,981	55,918	59,044	62,390	65,969	69,800
Selling Expenses	210	898	961	1,029	1,101	1,178	1,260	1,348	1,443	1,544
General Administrative Exp.	698	1,449	1,520	1,597	1,679	1,677	1,771	1,871	1,979	2,094
Total Product. Cost Before Int.	24,207	50,660	53,180	55,876	58,761	58,774	62,076	65,611	69,392	73,438
Interest										
Int. on Total Inv. Costs	5,267	5,267	5,267	5,267	4,514	3,762	3,009	2,257	1,504	752
Int. on Working Capital	1,843	1,228	614	0	0	0	0	0	0	0
Total Interest	7,110	6,496	5,881	5,267	4,514	3,762	3,009	2,257	1,504	752
Total Product. Cost Including Int.	31,318	57,156	59,062	61,143	63,276	62,536	65,086	67,868	70,897	74,191
(Unit Production Cost) (US¢/kg)	208.78	95.26	98.43	101.90	105.46	104.22	108.47	113.11	118.16	123.65

Table X-19-3 Unit Production Cost

(US¢/kg)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production (10^3 ton)	15	60	60	60	60	60	60	60	60	60
Variable Costs										
Raw Materials										
Cyclohexane	30.50	32.63	34.91	37.36	39.97	42.77	45.77	48.97	52.40	56.07
Sulfuric Acid	5.76	6.17	6.60	7.06	7.56	8.09	8.65	9.26	9.91	10.60
Ammonia	10.32	11.04	11.81	12.64	13.52	14.47	15.48	16.57	17.73	18.97
Catalysts and Chemicals	5.00	5.35	5.72	6.12	6.55	7.01	7.50	8.02	8.59	9.19
Ammonium Sulfate	-12.41	-13.27	-14.20	-15.20	-16.26	-17.40	-18.62	-19.92	-21.32	-22.81
Total Raw Material Costs	39.17	41.92	44.85	47.99	51.35	54.95	58.79	62.91	67.31	72.02
Utility										
Fuel	11.45	12.25	13.11	14.03	15.01	16.06	17.18	18.39	19.68	21.05
Total Utility Costs	11.45	12.25	13.11	14.03	15.01	16.06	17.18	18.39	19.68	21.05
Total Variable Costs	50.63	54.17	57.96	62.02	66.36	71.01	75.98	81.30	86.99	93.08
Wages	4.88	1.30	1.39	1.49	1.60	1.71	1.83	1.96	2.09	2.24
Depreciation										
Battery Limits	31.88	7.97	7.97	7.97	7.97	7.97	7.97	7.97	7.97	7.97
Off-Site	17.71	4.42	4.42	4.42	4.42	4.42	4.42	4.42	4.42	4.42
Building	0.92	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23	0.23
Royalty, Eng. Fee	8.96	2.24	2.24	2.24	2.24	0	0	0	0	0
Pre-Operating Expenses	2.02	0.50	0.50	0.50	0.50	0	0	0	0	0
Int. During Construction	8.91	2.22	2.22	2.22	2.22	0	0	0	0	0
Total Depreciation	70.42	17.60	17.60	17.60	17.60	12.63	12.63	12.63	12.63	12.63
Repairs and Insurance	24.49	6.12	6.12	6.12	6.12	6.12	6.12	6.12	6.12	6.12
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	4.88	1.30	1.39	1.49	1.60	1.71	1.83	1.96	2.09	2.24
Total Fixed Costs	104.69	26.34	26.52	26.72	26.93	22.18	22.42	22.67	22.95	23.24
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	155.32	80.52	84.49	88.75	93.30	93.19	98.40	103.98	109.94	116.33
Selling Expenses	1.40	1.49	1.60	1.71	1.83	1.96	2.10	2.24	2.40	2.57
General Administrative Exp.	4.65	2.41	2.53	2.66	2.79	2.79	2.95	3.11	3.29	3.49
Total Product. Cost Before Int.	161.38	84.43	88.63	93.12	97.93	97.95	103.46	109.35	115.65	122.39
Interest										
Int. on Total Inv. Costs	35.11	8.77	8.77	8.77	7.52	6.27	5.01	3.76	2.50	1.25
Int. on Working Capital	12.28	2.04	1.02	0	0	0	0	0	0	0
Total Interest	47.40	10.82	9.80	8.77	7.52	6.27	5.01	3.76	2.50	1.25
Total Product. Cost Including Int.	208.78	95.26	98.43	101.90	105.46	104.22	108.47	113.11	118.16	123.65

Table X-19-4 Sales Revenue

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Domestic										
Sales Price (US¢/kg)	100	107	114	122	131	140	150	160	171	184
Sales Amount (ton)	13,000	33,000	41,000	49,000	51,000	53,000	55,000	57,000	59,000	60,000
Sales Value	13,013	35,345	46,987	60,087	66,917	74,409	82,622	91,621	101,474	110,417
Export										
Sales Price (US¢/kg)	74	79	85	91	97	104	111	119	128	137
Sales Amount (ton)	2,000	27,000	19,000	11,000	9,000	7,000	5,000	3,000	1,000	0
Sales Value	1,492	21,551	16,227	10,052	8,800	7,324	5,597	3,593	1,281	0
Total Sales Value	14,505	56,897	63,215	70,139	75,718	81,733	88,220	95,214	102,756	110,417

Table X-19-5 Discounted Cash Flow Calculation

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Investment	123,889									
Land	1,747									
Inv. Excl. Land	122,142									
Total Sales Value	14,505	56,897	63,215	70,139	75,718	81,733	88,220	95,214	102,756	110,417
Total Product. Cost*	31,318	57,156	59,062	61,143	63,276	62,536	65,086	67,868	70,897	74,191
Income Before Tax	-16,813	-259	4,153	8,996	12,442	19,197	23,134	27,346	31,859	36,226
Income Tax	0	0	0	0	0	8,638	10,410	12,305	14,336	16,301
Income After Tax	-16,813	-259	4,153	8,996	12,442	10,558	12,723	15,040	17,522	19,924
Depreciation & Int.	17,673	17,059	16,444	15,830	15,078	11,340	10,588	9,835	9,083	8,331
Net Cash Flow (NCF)	860	16,800	20,597	24,826	27,520	21,898	23,311	24,875	26,605	28,255
Salvage Value (SV)	0	0	0	0	0	0	0	0	0	31,432
NCF + SV	860	16,800	20,597	24,826	27,520	21,898	23,311	24,875	26,605	59,687
Discount Rate	0.894587	0.800286	0.715925	0.640457	0.572945	0.512549	0.458520	0.410186	0.366947	0.328266
Present Value	769	13,444	14,745	15,899	15,767	11,223	10,688	10,203	9,762	19,593
(Total Present Value)										122,100
Internal Rate of Return (%)										11.7834

* Including Interest

Table X-20 Production Cost of Caprolactam (40,000.t/y)

Table X-20-1 Estimated Production Cost of Caprolactam

	Unit Consumption (kg/kg)	Price (US¢/kg)	Unit Cost (US¢/kg)
Variable Costs			
Raw Materials			
Cyclohexane	1.0	30.5	30.50
Sulfunic Acid	1.282	4.5	5.76
Ammonia	0.688	15.0	10.32
Catalysts and Chemicals			5.00
Ammonium Sulfate	1.7	-7.3	-12.41
Total Raw Material Costs			39.17
Utility			
Fuel	1.38	8.3	11.45
Total Utility Costs			11.45
Total Variable Costs			50.63
Wages			1.60
Depreciation			
Battery Limits			9.08
Off-Site			5.29
Building			0.31
Royalty, Eng. Fee			2.52
Pre-Operating Expenses			0.50
Int. During Construction			2.60
Total Depreciation			20.31
Repairs and Insurance			7.18
Taxes			0
Plant Overhead Costs			1.60
Total Fixed Costs			30.71
Running Royalty			0
Factory Costs			81.34
Selling Expenses			1.40
General Administrative Exp.			2.44
Total Product. Cost Before Int.			85.18
Interest			
Int. on Total Inv. Costs			5.13
Int. on Working Capital			3.25
Total Interest			8.39
Total Product. Cost Including Int.			93.58

Working Capital (10^3 US$)	10,846
Total Investment Cost*(10^3 US$)	95,355

* Including Land Price & Working Capital

Table X-20-2 Production Cost

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production (10^3 ton)	10	40	40	40	40	40	40	40	40	40
Variable Costs										
Raw Materials										
Cyclohexane	3,050	13,054	13,967	14,945	15,991	17,111	18,308	19,590	20,961	22,429
Sulfuric Acid	576	2,469	2,641	2,826	3,024	3,236	3,463	3,705	3,964	4,242
Ammonia	1,032	4,416	4,726	5,056	5,410	5,789	6,195	6,628	7,092	7,589
Catalysts and Chemicals	500	2,140	2,289	2,450	2,621	2,805	3,001	3,211	3,436	3,676
Ammonium Sulfate	-1,241	-5,311	-5,683	-6,081	-6,506	-6,962	-7,449	-7,971	-8,529	-9,126
Total Raw Material Costs	3,917	16,768	17,942	19,198	20,542	21,980	23,518	25,165	26,926	28,811
Utility										
Fuel	1,145	4,902	5,245	5,612	6,005	6,425	6,875	7,357	7,872	8,423
Total Utility Costs	1,145	4,902	5,245	5,612	6,005	6,425	6,875	7,357	7,872	8,423
Total Variable Costs	5,063	21,670	23,187	24,811	26,547	28,406	30,394	32,522	34,798	37,234
Wages	643	688	736	787	842	901	964	1,032	1,104	1,182
Depreciation										
Battery Limits	3,632	3,632	3,632	3,632	3,632	3,632	3,632	3,632	3,632	3,632
Off-Site	2,116	2,116	2,116	2,116	2,116	2,116	2,116	2,116	2,116	2,116
Building	124	124	124	124	124	124	124	124	124	124
Roayal. Eng. Fee	1,011	1,011	1,011	1,011	1,011	0	0	0	0	0
Pre-Operating Expenses	202	202	202	202	282	0	0	0	0	0
Int. During Construction	1,040	1,040	1,040	11,040	1,040	0	0	0	0	0
Total Depreciation	8,127	8,127	8,127	8,127	8,127	5,873	5,873	5,873	5,873	5,873
Repairs and Insurance	2,872	2,872	2,872	2,872	2,872	2,872	2,872	2,872	2,872	2,872
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	643	688	736	787	842	901	964	1,032	1,104	1,182
Total Fixed Costs	12,286	12,376	12,472	12,575	12,685	10,549	10,675	10,810	10,955	11,109
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	17,349	34,047	35,660	37,386	39,233	38,955	41,070	43,332	45,754	48,344
Selling Expenses	140	599	641	686	734	785	840	899	962	1,029
General Administrative Exp.	520	1,021	1,069	1,121	1,177	1,168	1,232	1,299	1,372	1,450
Total Product. Cost Before Int.	18,010	35,667	37,371	39,194	41,144	40,909	43,142	45,532	48,088	50,824
Interest										
Int. on Total Inv. Costs	4,110	4,110	4,110	4,110	3,523	2,935	2,348	1,761	1,174	587
Int. on Working Capital	1,301	867	433	0	0	0	0	0	0	0
Total Interest	5,411	4,978	4,544	4,110	3,523	2,935	2,348	1,761	1,174	587
Total Product. Cost Including Int.	23,421	40,645	41,915	43,304	44,668	43,845	45,491	47,293	49,263	51,411
(Unit Production Cost) (US¢/kg)	234.21	101.61	104.78	108.26	111.67	109.61	113.72	118.23	123.15	128.52

Table X-20-3 Unit Production Cost

(US¢/kg)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production (10^3 ton)	10	40	40	40	40	40	40	40	40	40
Variable Costs										
Raw Materials										
Cyclohexane	30.50	32.63	34.91	37.36	39.97	42.77	45.77	48.97	52.40	56.07
Sulfuric Acid	5.76	6.17	6.60	7.06	7.56	8.09	8.65	9.26	9.91	10.60
Ammonia	10.32	11.04	11.81	12.64	13.52	14.47	15.48	16.57	17.73	18.97
Catalysts and Chemicals	5.00	5.35	5.72	6.12	6.55	7.01	7.50	8.02	8.59	9.19
Ammonium Sulfate	-12.41	-13.27	-14.20	-15.20	-16.26	-17.40	-18.62	-19.92	-21.32	-22.81
Total Raw Material Costs	39.17	41.92	44.85	47.99	51.35	54.95	58.79	62.91	67.31	72.02
Utility										
Fuel	11.45	12.25	13.11	14.03	15.01	16.06	17.18	18.39	19.68	21.05
Total Utility Costs	11.45	12.25	13.11	14.03	15.01	16.06	17.18	18.39	19.68	21.05
Total Variable Costs	50.63	54.17	57.96	62.02	66.36	71.01	75.98	81.30	86.99	93.08
Wages	6.43	1.72	1.84	1.96	2.10	2.25	2.41	2.58	2.76	2.95
Depreciation										
Battery Limits	36.32	9.08	9.08	9.08	9.08	9.08	9.08	9.08	9.08	9.08
Off-Site	21.16	5.29	5.29	5.29	5.29	5.29	5.29	5.29	5.29	5.29
Building	1.24	0.31	0.31	0.31	0.31	0.31	0.31	0.31	0.31	0.31
Royalty, Eng. Fee	10.11	2.52	2.52	2.52	2.52	0	0	0	0	0
Pre-Operating Expenses	2.02	0.50	0.50	0.50	0.50	0	0	0	0	0
Int. During Construction	10.40	2.60	2 60	2.60	2.60	0	0	0	0	0
Total Depreciation	81.27	20.31	20.31	20.31	20.31	14.68	14.68	14.68	14.68	14.68
Repairs and Insurance	28.72	7.18	7.18	7.18	7.18	7.18	7.18	7.18	7.18	7.18
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	6.43	1.72	1.84	1.96	2.10	2.25	2.41	2.58	2.76	2.95
Total Fixed Costs	122.86	30.94	31.18	31.43	31.71	26.37	26.68	27.02	27.38	27.77
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	173.49	85.11	89.15	93.46	98.08	97.38	102.67	108.33	114.38	120.86
Selling Expenses	1.40	1.49	1.60	1.71	1.83	1.96	2.10	2.24	2.40	2.57
General Aministrative Exp.	5.20	2.55	2.67	2.80	2.94	2.92	3.08	3.24	3.43	3.62
Total Product. Cost Before Int.	180.10	89.16	93.42	97.98	102.86	102.27	107.85	113.83	120.22	127.06
Interest										
Int. on Total Inv. Costs	41.10	10.27	10.27	10.27	8.80	7.33	5.87	4.40	2.93	1.46
Int. on Working Capital	13.01	2.16	1.08	0	0	0	0	0	0	00
Total Interest	54.11	12.44	11.36	10.27	8.80	7.33	5.87	4.40	2.93	1.46
Total Product. Cost Including Int.	234.21	101.61	104.78	108.26	111.66	109.61	113.72	118.23	123.15	128.52

Table X-20-4 Sales Revenue

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Domestic										
Sales Price (US¢/kg)	100	107	114	122	131	140	150	160	171	184
Sales Amount (ton)	10,000	33,000	40,000	40,000	40,000	40,000	40,000	40,000	40,000	40,000
Sales Value	10,010	35,345	45,841	49,050	52,484	56,158	60,089	64,295	68,796	73,611
Export										
Sales Price (US¢/kg)	74	79	85	91	97	104	111	119	128	137
Sales Amount (ton)	0	7,000	0	0	0	0	0	0	0	0
Sales Value	0	5,587	0	0	0	0	0	0	0	0
Total Sales Value	10,010	40,932	45,841	49,050	52,484	56,158	60,089	64,295	68,796	73,611

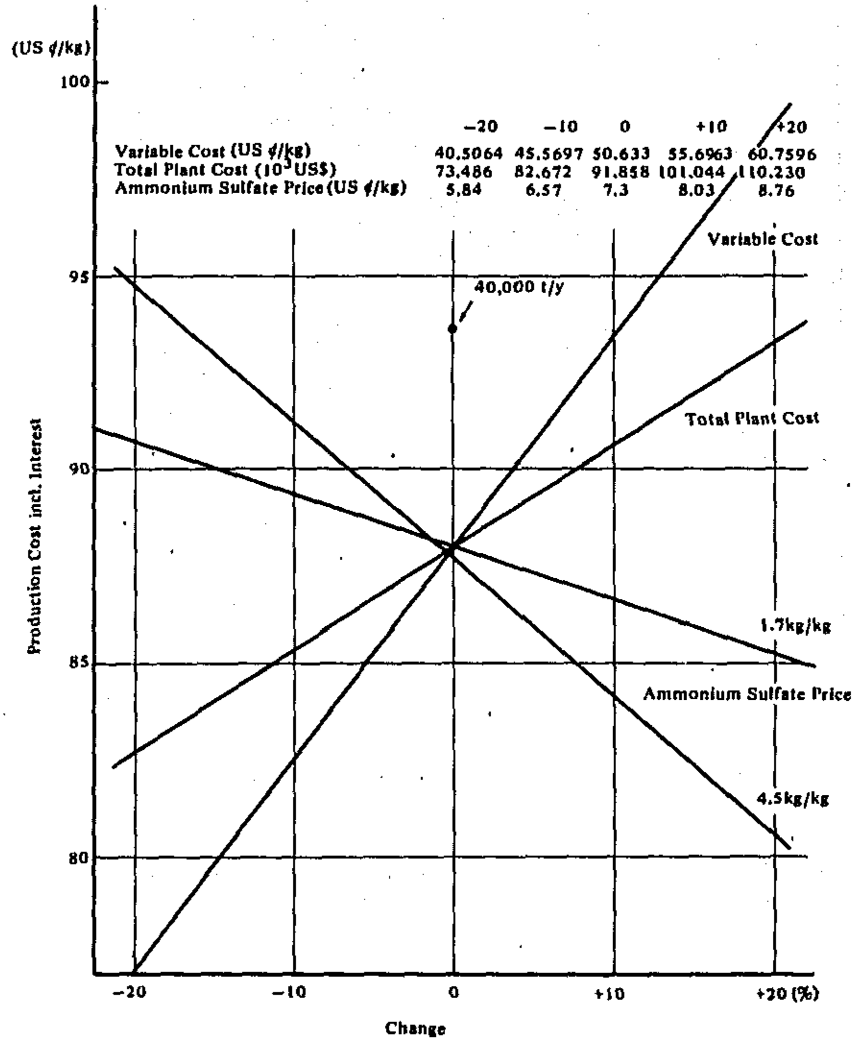
Table X-20-5 Discounted Cash Flow Calculation

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Investment	95,355									
Land	1,425									
Inv. Excl. Land	93,930									
Total Sales Value	10,010	40,932	45,841	49,050	52,484	56,158	60,089	64,295	68,796	73,611
Total Product. Cost*	23,421	40,645	41,915	43,304	44,668	43,845	45,491	47,293	49,263	51,411
Income Before Tax	13,411	287	3,926	5,746	7,816	12,313	14,598	17,002	19,533	22,200
Income Tax	0	0	0	0	0	5,540	6,569	7,650	8,789	9,990
Income After Tax	13,411	287	3,926	5,746	7,816	6,772	8,028	9,351	10,743	12,210
Depreciation & Int.	13,539	13,105	12,672	12,238	11,650	8,809	8,221	7,634	7,047	6,460
Net Cash Flow (NCF)	128	13,392	16,598	17,984	19,466	15,581	16,249	16,985	17,790	18,670
Salvage Value (SV)	0	0	0	0	0	0	0	0	0	23,924
NCF + SV	128	13,392	16,598	17,984	19,466	15,581	16,249	16,985	17,790	42,594
Discount Rate	0.904033	0.817276	0.738844	0.667939	0.603839	0.545890	0.493503	0.446143	0.403328	0.364622
Present Value	115	10,944	12,263	12,012	11,754	8,505	8,019	7,577	7,175	15,530
(Total Present Value)										93,899
Internal Rate of Return (%)										10.6154

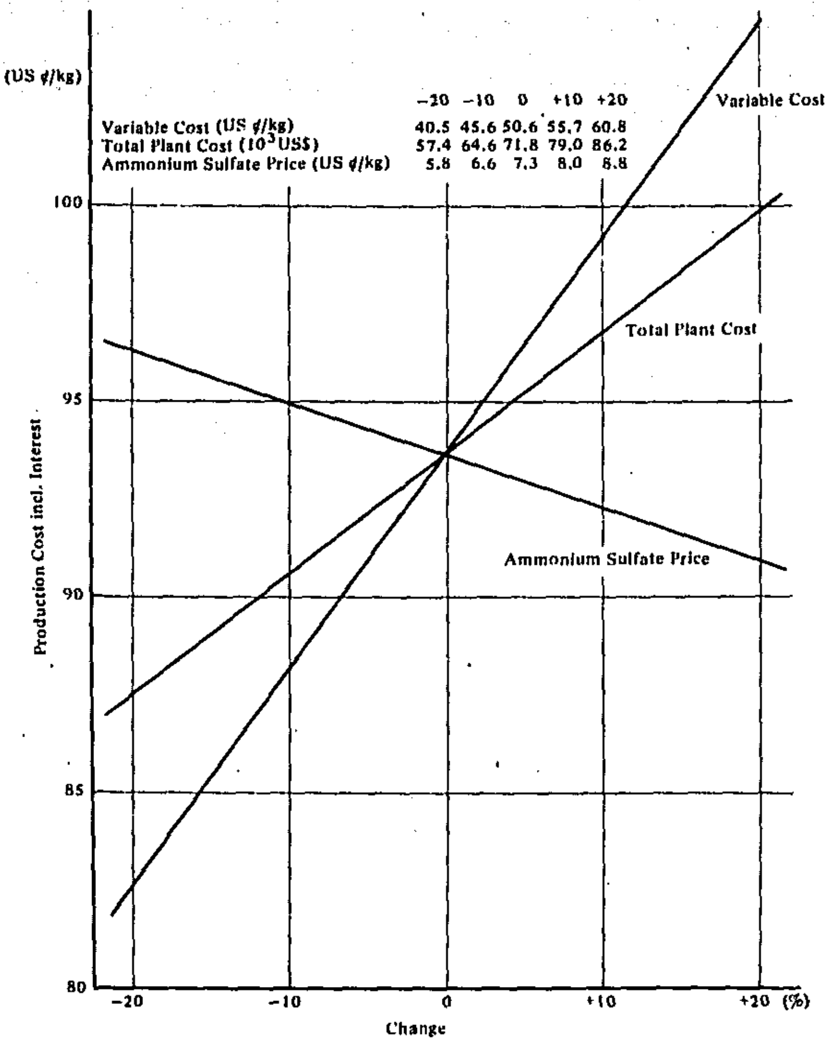
* Including Interest

Figures X-13 and X-14 show the variation which takes place in the production cost when the variable cost, total plant cost and ammonium sulfate cost individually fluctuated by ±20% regarding both 60,000 t/y and 40,000 t/y operations. The effect of the variable cost upon the production cost is the most conspicuous so that a variation by 10% in the variable cost corresponds to a 20% variation in the total plant cost.



	−20	−10	0	+10	+20
Variable Cost (US ¢/kg)	40.5064	45.5697	50.633	55.6963	60.7596
Total Plant Cost (10^3 US$)	73.486	82.672	91.858	101.044	110.230
Ammonium Sulfate Price (US ¢/kg)	5.84	6.57	7.3	8.03	8.76

Figure X-13 Sensitivity Analysis of Caprolactam Production Cost (60,000 t/y)



Further, the variable costs obtained by varying the extent of ammonium sulfate by production from 1.6 to 4.5 kg, there will be little change in the variable cost as shown in Table X-21. However, in practice, it is not possible to undertake such a wide range of variation of the ammonium sulfate by-production extent without changing the facilities. Therefore, although little change will be caused in the variable cost, the fixed cost such as plant cost, warehouse cost, bagging machinery cost, etc. will increase. Also, when the extent of ammonium sulfate by-production increases, the fluctuation in the ammonium sulfate market price will strongly affect the production cost of caprolactam. (Refer to Figure X-13.) The fluctuation in ammonium sulfate price is much greater than the possible variation in the construction cost and the raw materials cost. This signifies that a high extent of ammonium sulfate by-production is not desirable in view of the stability in the operation on enterprise.

Table X-21 Ammonium Sulfate By-Production Amount and Variable Cost in Caprolactam Production (60,000 t/y)

(/kg Caprolactam)

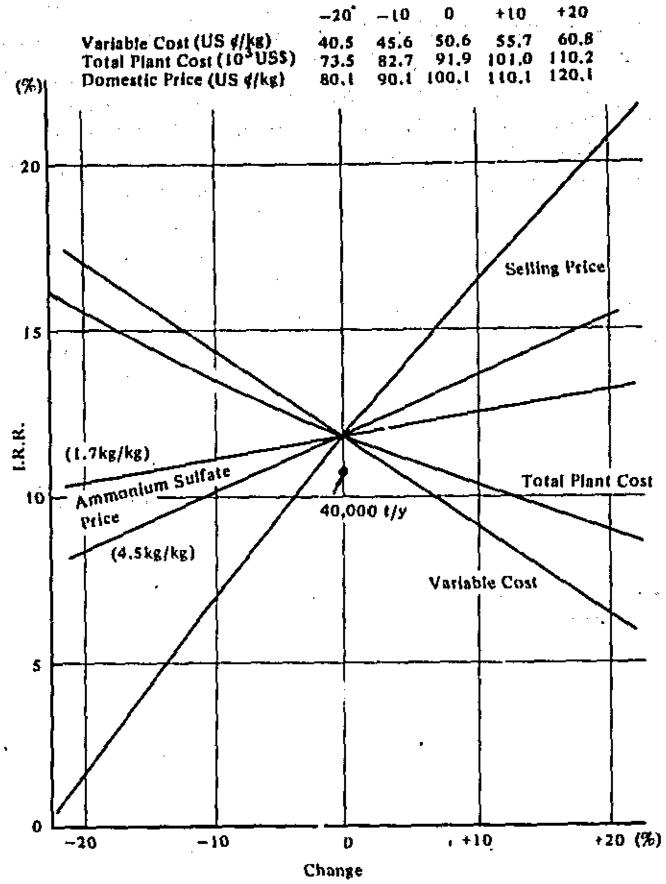
By-Produced Ammonium Sulfate (kg)	1.6	1.7	2.0	2.6	4.5
H_2SO_4 Consumption (kg)	1.208	1.282	1.505	1.950	3.361
NH_3 " (")	0.662	0.688	0.765	0.920	1.409
Variable Cost (US¢/kg)	50.6	50.6	50.6	50.5	50.4

5-4 Evaluation by the DCF Method

Figures X-15 and X-16 show the variation which will take place in the IRR if ±20% each variation should take place in the market price, total plant cost, variable cost, and the price and the quantity of by-produced ammonium sulfate.

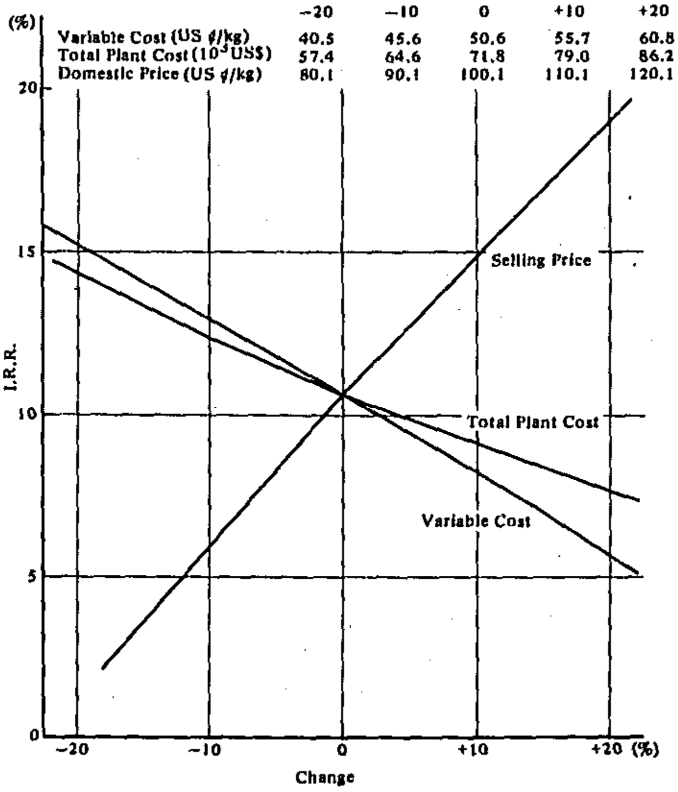
As far as the standard case is concerned, the IRR is 11.8% for 60,000 t/y and 10.6% for 40,000 t/y by showing an advantage in the case of the former although the difference is slight. In view of the production cost, the 40,000 t/y operation is higher by US¢5.6/kg when compared with the 60,000 t/y, however, the effect of the exportation rate made the IRR difference extremely small. In the case of the 60,000 t/y, a considerable extent of exportation must be carried out until the domestic demand approaches the production amount, thereby lowering the IRR level. The market price is the major factor which exerts the strongest influence on the IRR, followed by a variable cost and the total plant cost.

In the case of 60,000 t/y, the IRR is 11.8% and this rate is small if the target IRR rate is considered as 15%. In order to bring the IRR up to 15%, the average market price must be increased by 7% while the remaining total plant cost unchanged, or the variable cost must be reduced by 12%. The reduction in the variable cost by 12% corresponds to the reduction in the raw materials cost by US¢6/kg. Due to the fact that the prices of sulfuric acid, ammonia and ammonium sulfate will cancel one another, it is not possible to drastically lower the variable cost by means of these items. Therefore, it is necessary to reduce the variable cost by lowering the cyclohexane cost or the fuel cost. At present,



	−20	−10	0	+10	+20
Variable Cost (US ¢/kg)	40.5	45.6	50.6	55.7	60.8
Total Plant Cost (10^6 US$)	73.5	82.7	91.9	101.0	110.2
Domestic Price (US ¢/kg)	80.1	90.1	100.1	110.1	120.1

Figure X-15 Sensitivity Analysis of IRR in Caprolactam Production (60,000 t/y)



	−20	−10	0	+10	+20
Variable Cost (US ¢/kg)	40.5	45.6	50.6	55.7	60.8
Total Plant Cost (10^6 US$)	57.4	64.6	71.8	79.0	86.2
Domestic Price (US ¢/kg)	80.1	90.1	100.1	110.1	120.1

Figure X-16 Sensitivity Analysis of IRR in Caprolactam Production (60,000) the evaluation of the fuel cost is made on the basis of heavy oil, however, if it is feasible to employ a low-cost natural gas, it would be possible to realize a large extent of cost reduction. If the cost of natural gas is 1/2 of the heavy oil price, the variable cost will be reduced by 10%, thereby increasing the IRR level up to 14.5%. Also, when the employment of natural gas is undertaken, such a utilization will be particularly advantageous to the operation of the PNC process which requires a large extent of electricity.

The ammonium sulfate cost involves the highest extent of fluctuation amongst all the variable factors treated in the sensitivity analysis. If ammonium sulfate by-production extent is 1.7, the fluctuation in the ammonium sulfate price will exert a comparatively slight influence on the IRR. However, if the by-production increased to 4.5, the effect will be considerable. Therefore, high by-production of ammonium sulfate should be avoided as much as possible in view of the maintenance of stability in the operation of enterprise.

6. Extent of Contribution to the Indonesian National Economy

The extent of contribution to Indonesian national economy by the implementation of this project was assessed in view of the saving in foreign currency by employing the shadow pricing method. The results are shown in Tables X-22 and X-23.

Table X-22 Foreign Currency Saving Attainable by Caprolactam 60,000 t/y Production

(10^3 US$)

	A	B	C	D	E	F	Present Value
1977	14,767	2,957	5,261	1,865	3,954	4,460	3,878
1978	44,249	8,033	5,261	26,938	16,923	40,970	30,979
1979	55,887	10,679	5,261	20,283	18,108	42,122	27,695
1980	68,938	13,656	5,261	12,565	19,376	43,210	24,705
1981	76,209	15,208	19,995	11,000	20,732	31,274	15,548
1982	84,159	16,911	19,242	9,155	22,183	34,978	15,121
1983	92,849	18,777	18,491	6,996	23,736	38,841	14,601
1984	102,344	20,823	17,740	4,491	25,398	42,874	14,015
1985	112,712	23,062	16,988	1,601	27,176	47,087	13,385
1986	122,316	25,095	16,236	0	29,078	51,907	12,830
Total	774,430	155,201	129,736	94,894	206,664	377,723	172,757

Notes: A: Sales Value of Product
B: Export Value of Raw Materials
C: Repayment of Foreign Loan and Interest on Foreign Loan
D: Export Value of Product
E: Import Value of Sub-Raw Materials
F: $F = A - B - C + D - E$
Yearly Foreign Currency Saving
Present Value: $\frac{Fn}{(1+0.15)^n}$

The amount of saving attainable in foreign currency for a period of ten years from the implementation of the project was calculated on the basis of the shadow exchange rate of US$1 = Rp519. The result reveals, as shown in Table X-22, that the amount will be US$378 million in the case of producing 60,000 t/y of caprolactam. When discounted by 15%/y, the amount becomes US$173 million which implies a considerable contribution to the foreign currency saving of Indonesia.

Then, by employing the method described in the clause pertaining to the Survey Method, the economic evaluation in view of national economy was undertaken concerning the implementation of this project by utilizing the exchange rate of US$1 = Rp519 and the labor cost on the basis of 40% of the actually prevailing wages. The obtained results are shown in Table X-23. The results indicate that IRR equals to 14.5% which almost satisfies the standard target value.

It must be noted that the above calculations do not include the external effects exerted by the implementation of this project.

It is forecast that raw materials for synthetic fiber will suffer a serious shortage in the future and that the development of the synthetic fiber industry in Indonesia will be considerably impaired unless this project is implemented. Therefore, realization of this project seems indispensable to ensure the development of the synthetic fiber industry in Indonesia and to attain one of the objectives of the Second Five-year Development Plan of Indonesia, i.e., "The attainment of sufficient supply of food and clothing according to the purchasing power of the people".

Table X-23 Evaluation of National Benefit by Caprolactam Production

Table X-23-1 Estimated Production Cost of Caprolactam (60,000 t/y)

	Unit Consumption (kg/kg)	Price (US¢/kg)	Unit Cost (US¢/kg)
Variable Costs			
Raw Materials			
Cyclohexane	1.0	32.0	32.00
Sulfunic Acid	1.282	5.625	7.21
Ammonia	0.688	18.75	12.90
Catalysts and Chemicals			6.25
Ammonium Sulfate	1.7	-10.375	17.63
Total Raw Mterial Costs			40.72
Utility			
Fuel	1.38	10.375	14.31
Total Utility Costs			14.31
Total Variable Costs			55.04
Wages			0.48
Depreciation			
Battery Limits			9.83
Off-Site			5.36
Building			0.24
Royalty, Eng. Fee			2.80
Pre-Operating Expenses			0.55
Int. During Construction			2.70
Total Depreciation			21.49
Repairs and Insurance			7.44
Taxes			0
Plant Overhead Costs			0.48
Total Fixed Costs			29.92
Running Royalty			0
Factory Costs			84.96
Selling Expenses			1.40
General Administrative Exp.			2.54
Total Product. Cost Before Int.			88.91
Interest			
Int. on Total Inv. Costs			5.33
Int. on Working Capital			3.39
Total Interest			8.73
Total Product. Cost Including Int.			97.64

Working Capital (10^3 US$)	16,993
Total Investment Cost *(10^3 US$)	148,636

* Including Land Price & Working Capital

Table X-23-2 Production Cost

(10^3 US$/y)

Plant Life (Years) Year (Year)	1 1977	2 1978	3 1979	4 1980	5 1981	6 1982	7 1983	8 1984	9 1985	10 1986
Stream Factor	0.25	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Annual Production (10^3 ton)	15	60	60	60	60	60	60	60	60	60
Variable Costs										
Raw Materials										
Cyclohexane	4,800	20,544	21,982	23,520	25,167	26,928	28,814	30,831	32,989	35,298
Sulfuric Acid	1,081	4,629	4,953	5,300	5,671	6,068	6,493	6,947	7,434	7,954
Ammonia	1,935	8,281	8,861	9,481	10,145	10,855	11,615	12,428	13,298	14,229
Catalysts and Chemicals	937	4,012	4,293	4,593	4,915	5,259	5,627	6,021	6,443	6,894
Ammonium Sulfate	-2,645	-11,323	-12,115	-12,964	-13,871	-14,842	-15,881	-16,993	-18,182	-9,455
Total Raw Material Costs	6,108	26,144	27,974	29,933	32,028	34,270	36,669	39,236	41,982	44,921
Utility										
Fuel	2,147	9,191	9,835	10,523	11,260	12,048	12,892	13,794	14,760	15,793
Total Utility Costs	2,147	9,191	9,835	10,523	11,260	12,048	12,892	13,794	14,760	15,793
Total Variable Costs	8,256	35,336	37,810	40,456	43,285	46,318	49,561	53,030	56,742	60,714
Wages	293	313	335	358	384	410	439	470	503	538
Depreciation										
Battery Limits	5,901	5,901	5,901	5,901	5,901	5,901	5,901	5,901	5,901	5,901
Off-Site	3,217	3,217	3,217	3,217	3,217	3,217	3,217	3,217	3,217	3,217
Building	148	148	148	148	148	148	148	148	148	148
Royalty, Eng. Fee	1,680	1,680	1,680	1,680	1,680	0	0	0	0	0
Pre-Operating Expenses	330	330	330	330	330	0	0	0	0	0
Int. During Construction	1,621	1,621	1,621	1,621	1,621	0	0	0	0	0
Total Depreciation	12,899	12,899	12,899	12,899	12,899	9,267	9,267	9,267	9,267	9,267
Repairs and Insurance	4,469	4,469	4,469	4,469	4,469	4,469	4,469	4,469	4,469	4,469
Taxes	0	0	0	0	0	0	0	0	0	0
Plant Overhead Costs	293	313	335	358	384	410	439	470	503	538
Total Fixed Costs	17,955	17,996	18,040	18,087	18,137	14,559	14,616	14,678	14,744	14,814
Running Royalty	0	0	0	0	0	0	0	0	0	0
Factory Costs	26,211	53,332	55,850	58,543	61,426	60,878	64,178	67,708	71,486	75,529
Selling Expenses	210	898	961	1,029	1,101	1,178	1,260	1,348	1,443	1,544
General Administrative Exp.	786	1,599	1,675	1,756	1,842	1,826	1,925	2,031	2,144	2,265
Total Product. Cost Before Int.	27,207	55,831	58,487	61,329	64,370	63,882	67,364	71,089	75,075	79,339
Interest										
Int. on Total Inv. Costs	6,398	6,398	6,398	6,398	5,484	4,570	3,656	2,742	1,828	914
Int. on Working Capital	2,039	1,359	679	0	0	0	0	0	0	0
Total Interest	8,438	7,758	7,078	6,398	5,484	4,570	3,656	2,742	1,828	914
Total Product. Cost Including Int.	35,646	63,590	65,566	67,728	69,854	68,453	71,020	73,831	76,903	80,253
(Unit Production Cost) US¢/kg	237.64	105.98	109.27	112.88	116.42	114.08	118.36	123.05	128.17	133.75

Table X-23-3 Sales Revenue

(10^3 US$/y)

Plant Life (Years)	1	2	3	4	5	6	7	8	9	10
Year (Year)	1977	1978	1979	1980	1981	1982	1983	1984	1985	1986
Domestic										
Sales Price (US¢/kg)	107	115	123	131	141	151	161	173	185	198
Sales Amount (ton)	13,000	33,000	41,000	49,000	51,000	53,000	55,000	57,000	59,000	60,000
Sales Value	14,007	38,046	50,578	64,679	72,031	80,096	88,937	98,623	109,229	118,856
Export										
Sales Price (US¢/kg)	93	99	106	114	122	130	139	149	160	171
Sales Amount (ton)	2,000	27,000	19,000	11,000	9,000	7,000	5,000	3,000	1,000	0
Sales Value	1,865	26,939	20,284	12,565	11,000	9,155	6,997	4,492	1,602	0
Total Sales Value	15,872	64,986	70,863	77,245	83,032	89,251	95,934	103,115	110,831	118,856

Table X-23-4 Discounted Cash Flow Calculation

(10³ US$/y)

Plant Life (Years)	1	2	3	4	5	6	7	8	9	10
Year (Year)	1977	1978	1979	1980	1981	1982	1983	1984	1985	1986
Investment	148,636									
Land	1,747									
Inv. Excl. Land	146,889									
Total Sales Value	15,872	64,986	70,863	77,245	83,032	89,251	95,934	103,115	110,831	118,856
Total Product. Cost*	35,646	63,590	65,566	67,728	69,854	68,453	71,020	73,831	76,903	80,253
Income Before Tax	-19,774	1,396	5,297	9,517	13,178	20,798	24,914	29,284	33,928	38,603
Income Tax	0	0	0	0	0	0	0	0	0	0
Income After Tax	-19,774	1,396	5,297	9,517	13,178	20,798	24,914	29,284	33,928	38,603
Depreciation & Int.	21,337	20,658	19,978	19,298	18,384	13,838	12,924	12,010	11,096	10,182
Net Cash Flow (NCF)	1,563	22,054	25,275	28,815	31,562	34,636	37,838	41,294	45,024	48,785
Salvage Value (SV)	0	0	0	0	0	0	0	0	0	36,050
NCF + SV	1,563	22,054	25,275	28,815	31,562	34,636	37,838	41,294	45,024	84,835
Discount Rate	0.873141	0.762375	0.665661	0.581216	0.507484	0.443105	0.386893	0.337812	0.294958	0.257540
Present Value	1,364	16,813	16,824	16,747	16,017	15,347	14,639	13,949	13,280	21,848
(Total Present Value)										146,832
Internal Rate of Return (%)										14.529

* Including Interest