CHAPTER IV
RESULTS AND DISCUSSION

This chapter discusses on the activity testing of $MgHPO_4/\gamma\text{-}Al_2O_3$ catalyst for ethylene production from bio-ethanol and the economic pre-feasibility study on the commercial plants for the catalytic dehydration of bio-ethanol to ethylene.

4.1 Catalyst Testing for Catalytic Dehydration of Bio-ethanol to Ethylene

This section discusses about the investigation of 0.5 wt% $MgHPO_4/\gamma\text{-}Al_2O_3$ catalyst for its ethanol conversion and ethylene selectivity on the catalytic dehydration of bio-ethanol. The effects of reaction temperature and concentration of bio-ethanol in the feed are discussed as follows.

4.1.1 Effect of Reaction Temperature

As we know, the catalytic dehydration of ethanol to ethylene is highly endothermic reaction, and requires a large amount of energy to maintain activity level. Therefore, the ethanol conversion and product distribution strongly depends on reaction temperature. The ethanol conversion and the selectivity of ethylene over the 0.5 wt% $MgHPO_4/\gamma\text{-}Al_2O_3$ catalyst with the temperature in the range of 370 °C to 460 °C are shown in Figure 4.1. It is observed that the ethanol conversion and the selectivity of ethylene increase with increasing reaction temperature, whereas the selectivity of diethyl ether (DEE) decreases. This may be due to the reason that the ethylene formation occurs via intra-molecular dehydration at a high reaction temperature, while diethyl-ether can be produced via inter-molecular dehydration at low temperatures. Furthermore, other by-products, which are methane, ethane, propylene, C_4-C_5 and carbon dioxide, are also formed in the catalytic dehydration of ethanol as shown in Table 4.1. It can be observed that the increasing reaction temperature results in an increase of most by-products, excepting C_4-C_5, which is in contrary. The results reveal that a high reaction temperature can promote by-products by cracking and oligermerization of ethylene, and other side reactions of ethanol dehydration, leading to an increase in methane, ethane, propylene, and carbon dioxide and a decrease in C_4-C_5. More than 95% ethanol conversion and 90% ethylene selectivity can be obtained at reaction temperatures above 400 °C. The ethanol conversion is nearly 99% at the temperature of 460 °C, and ethylene selectivity is as high as 93.5% at the temperature of 430 °C. However, the highest yield of ethylene (97.1% conversion and 93.5% ethylene selectivity) is obtained at temperature of 430 °C. It can be suggested that a high reaction temperature favors high ethanol conversion and high ethylene selectivity, whereas the selectivity of diethyl ether is suppressed.

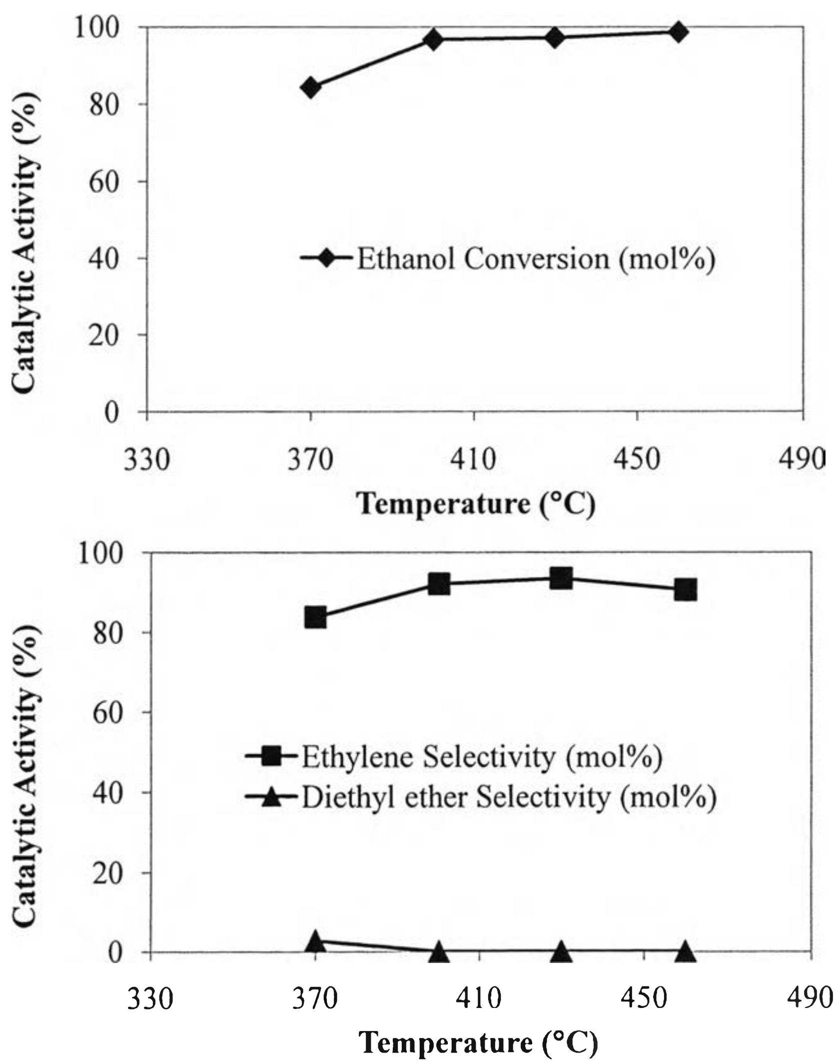


Figure 4.1 Ethanol conversion, ethylene selectivity, and diethyl ether selectivity over 0.5 wt% $MgHPO_4/\gamma$-Al_2O_3 (LHSV = 1.0 h^{-1}, 99.5% ethanol feed).

Table 4.1 Product distribution at various reaction temperatures over 0.5 wt% $MgHPO_4/\gamma$-Al_2O_3 (LHSV = 1.0 h^{-1}, 99.5% ethanol feed)

Temperature (°C)	X_{EtOH} (mol%)	Selectivity (mol%)						
		C_2H_4	C_2H_6	C_3H_6	C_4-C_5	CH_4	CO_2	DEE
370	84.2	83.8	0.55	0.32	12.6	0	0.05	2.67
400	96.6	92.1	0.55	0.75	6.33	0.03	0.25	0
430	97.1	93.5	0.51	0.79	4.68	0	0.53	0
460	98.6	90.6	0.66	1.34	6.16	0.17	1.05	0

According to the results, the reaction at a temperature of 370 °C tends to give a low ethanol conversion (84.2%) and low ethylene selectivity (83.8%), but it gives a high C_4-C_5 selectivity (12.6%). However, C_4-C_5 intermediates are further cracked to by-products such as methane and carbon dioxide at a higher temperature. No liquid hydrocarbons are found at any temperature. With increasing reaction temperature, the ethylene selectivity significantly increases, which may be resulted from the cracking reaction. However, the ethylene selectivity at a temperature of 460 °C decreases, whereas by-products (methane, ethane, propylene, and carbon dioxide) increase as shown in Table 4.1. This may be due to ethylene cracking ability and side reactions of ethanol dehydration that are promoted at a high temperature as mentioned above, leading to a decrease in ethylene selectivity at a temperature of 460 °C.

Additionally, the increasing reaction temperature slightly increases the coke formation on the catalyst as observed in Table 4.2. It can be suggested that higher reaction temperature causes higher amount of coke formation on the catalyst.

Table 4.2 Coke formation on spent catalysts at various reaction temperatures

Temperature (°C)	Coke formation (wt%)
370	2.1
400	2.1
430	2.3
460	2.5

4.1.2 Effect of Bio-ethanol Feed Concentration

The effects of bio-ethanol concentrations (40-50%, 95%, and 99.5%) in the feed on the ethanol conversion and the ethylene selectivity over the 0.5 wt% $MgHPO_4/\gamma\text{-}Al_2O_3$ catalyst are shown in Figure 4.2. It can be observed that at a fixed temperature the ethanol conversion decreases with decreasing ethanol concentration. When the temperatures are lower than 400 °C, the ethanol conversion significantly decreases. Especially, the ethanol conversion from using 40-50% ethanol as the feed rapidly decreases. This may be due to the competitive adsorption of water and ethanol on the active sites of catalyst surface and the decrease in bed temperature, that make ethanol be hardly converted to ethylene. However, the effect of water in the feed is insignificant when the reaction temperatures are higher than 400 °C. The ethylene selectivity significantly increases with decreasing ethanol concentration in the temperature above 400 °C, whereas the selectivity of diethyl ether is in the opposite. It can be suggested that the water in the feed has a positive effect on the ethanol conversion and ethylene selectivity in the temperature regions above 400 °C.

At the reaction temperature of 460 °C, the product distribution with various ethanol concentrations is shown in Table 4.3. It can be seen that ethanol conversion decreases with decreasing ethanol concentration in the feed due to the reasons mentioned above, whereas the ethylene selectivity increases with decreasing ethanol concentration. The ethylene selectivity is as high as 94.8%, when 40-50% ethanol is used as the feed. Furthermore, it also can be observed that a lower ethanol concentration results in a decrease in the production of by-products (methane, ethane, propylene, C_4-C_5), except that of CO_2, which is on contrary.

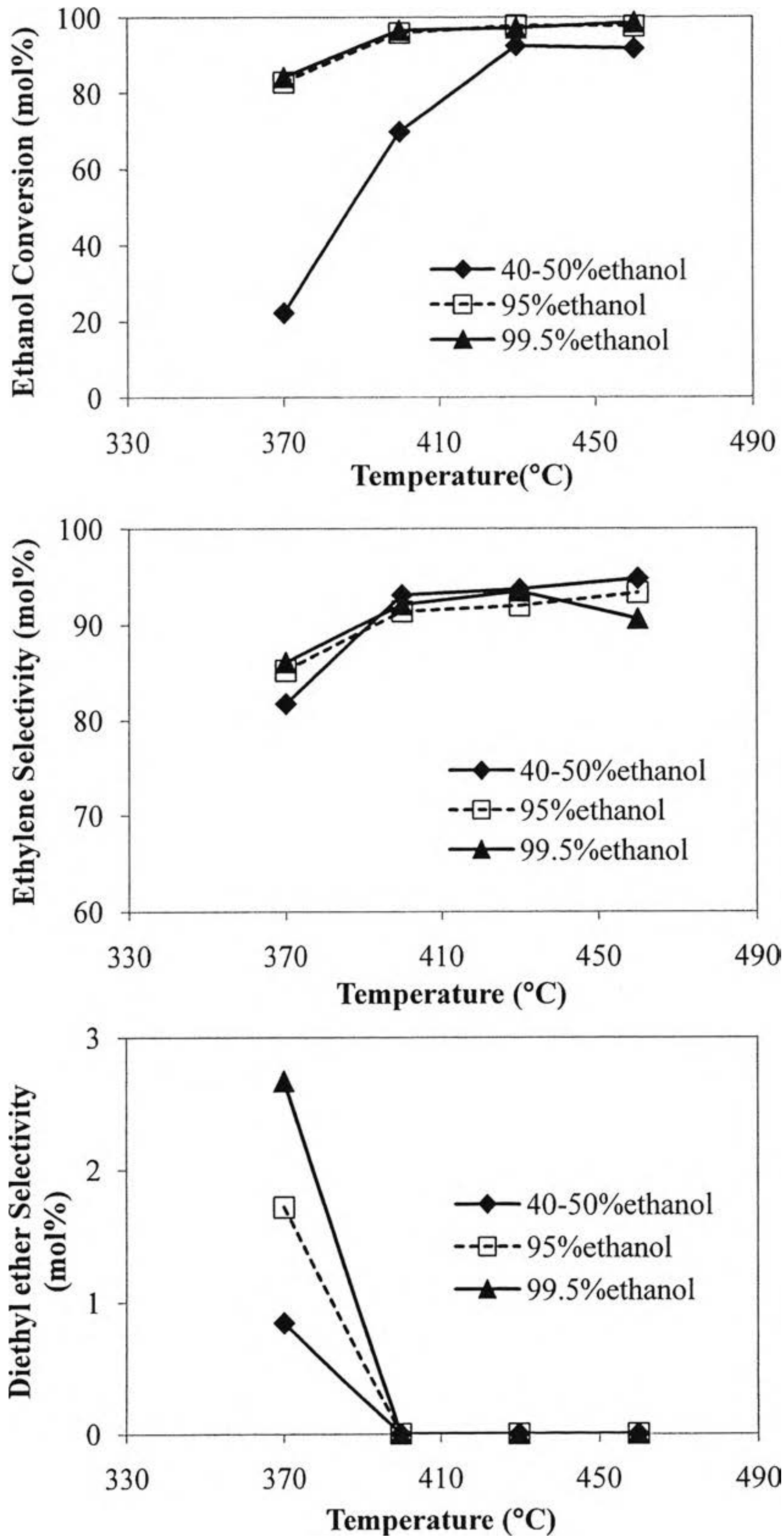


Figure 4.2 Ethanol conversion, ethylene selectivity, and diethyl ether selectivity over 0.5 wt% $MgHPO_4/\gamma$-Al_2O_3 (LHSV = 1.0 h^{-1}).

Table 4.3 Product distribution at various ethanol concentrations over 0.5 wt% $MgHPO_4/\gamma$-Al_2O_3 (LHSV = 1.0 h^{-1}, temperature of 460 °C)

% Ethanol	X_{EtOH} (mol%)	Selectivity (mol%)						
		C_2H_4	C_2H_6	C_3H_6	C_4-C_5	CH_4	CO_2	DEE
40-50	91.7	94.8	0.15	0.74	2.47	0.05	1.79	0
95	97.6	93.3	0.44	1.01	4.22	0.08	0.92	0
99.5	98.6	90.6	0.66	1.34	6.16	0.17	1.05	0

The results reveal that the use of low ethanol concentration (40-50% ethanol) is a promising way for the catalytic dehydration of ethanol to ethylene. However, the reactor must be carefully designed to be able to handle supersaturated steam at 460 °C, and the stability of catalyst needs to be further investigated.

From all of catalytic activity results, it can be seen that the operating conditions giving the highest yield of ethylene (97.6% ethanol conversion and 93.3% ethylene selectivity) is at the reaction temperature of 460 °C with using 95% ethanol as the feed. As compared with the commercial catalyst (**SYNDOL**), 0.5 wt% $MgHPO_4/\gamma$-Al_2O_3 catalyst gives the higher yield at 91.1%, whereas the commercial catalyst gives 96.0%. It can be seen that the difference is 4.97%, which is acceptable for verification. However, the difference may be resulted from what was added in the commercial catalyst as an additive. Therefore, 0.5 wt% $MgHPO_4/\gamma$-Al_2O_3 catalyst should be further studied to maximize the yield of ethylene for commercial use.

4.2 Pre-Feasibility Study on Commercial Plants for Catalytic Dehydration of Bi-ethanol to Ethylene

4.2.1 Background, Objective and Scope of Work

The technology background and details for catalytic dehydration of bio-ethanol to ethylene process is based on the polymer-grade ethylene production from the processes of Chematur Engineering AB and Petrobras. The difference between two processes is the adiabatic reaction system, which is described in details in Section *4.2.2.2.* The plants were designed to manufacture polymer-grade ethylene from bio-ethanol via catalytic dehydration. The ethylene production capacity is 33,000 metric tons per year for Chematur process, and 200,000 metric tons per year for Petrobras process, using a commercial grade ethanol as a raw material. The block diagram of two processes is shown in Figures 4.3 and 4.4, respectively.

The objective of this work was to perform the economic evaluation of the two commercial plants for polymer-grade ethylene production from bio-ethanol.

Scope and limitations of this work cover as follows:

(a) The raw material is 95% ethanol for Chematur plant and 93% ethanol for Petrobras plant, supplied by ethanol production companies via trucks. Ethanol storage is included in the design.

(b) The ethylene product is stored in the refrigerated ethylene tank, so a refrigerated ethylene storage is included in the design.

(c) The transportation of ethylene product to customers is done by a pipeline system.

(d) The steam used in the process can be bought. So, no steam generator is included in the design.

(e) Electricity is purchased from Provincial Electricity Authority.

(f) Waste water from the plant is sent to a waste water treatment system located on-site. Therefore, a waste water treatment system is included in the design.

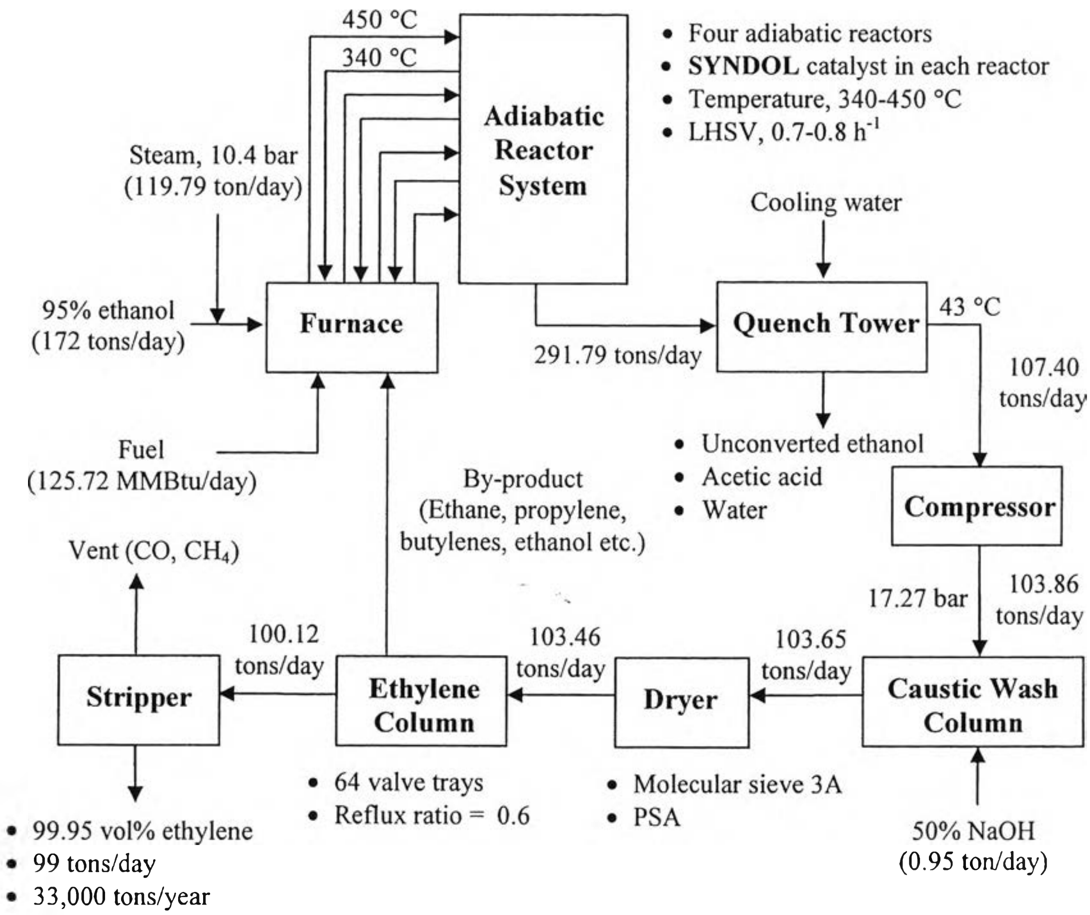


Figure 4.3 Block diagram of polymer-grade ethylene production via catalytic dehydration from bio-ethanol based on Chematur process.

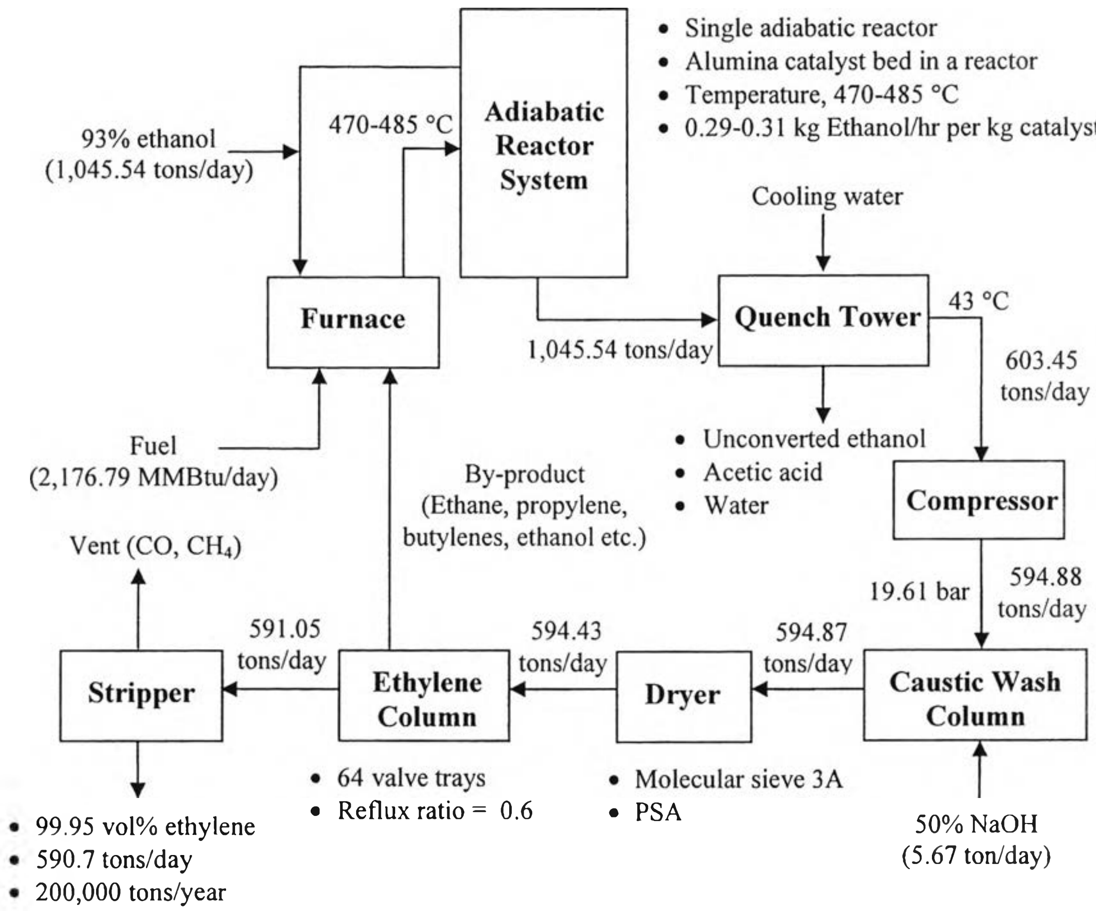


Figure 4.4 Block diagram of polymer-grade ethylene production via catalytic dehydration from bio-ethanol based on Petrobras process.

4.2.2 Process Description

4.2.2.1 General Plant Description

A polymer-grade ethylene is produced via the catalytic dehydration of bio-ethanol. Bio-ethanol used as a feed is vaporized and heated in a furnace where its temperature is raised to a reaction temperature before being fed into an adiabatic reactor system packed with a catalyst in the reactor. At the adiabatic reactor, ethanol is converted to ethylene when ethanol contacts with the catalyst. The effluent stream leaving from the adiabatic reactor system is passed to the quench tower where water and condensable polar-substances such as unconverted ethanol and acetic acid are removed. Subsequently, the crude ethylene leaving the top of quench tower is compressed, and passed through a caustic wash column and then a fixed bed gas drying system in order to reduce the concentrations of carbon dioxide and water to specification levels. With the aim for producing polymer-grade ethylene (99.95 vol%), the crude ethylene is sent to an ethylene column and a stripper in order to remove heavy and light contaminants. Based on the overview processes in Figures 4.3 and 4.4, the process description of two processes can be summarized and listed in Tables 4.4 and 4.5, respectively.

Table 4.4 Summary of plant description for ethanol dehydration to ethylene based on Chematur process

Description	Value
Plant Capacity (based on ethylene product)	33,000 tons/year 99 tons/day 4.125 tons/hour
Working time (hours/year)	8,000 hours/year ; (333 days/year)
Number of adiabatic reactors	4 adiabatic reactors
Catalyst	**SYNDOL** catalyst (99% conversion and 97% ethylene selectivity)
Specification of ethanol as feed	95% ethanol (hydrous ethanol)
Operating temperature	340-450 °C
LHSV (liquid hourly space velocity)	0.7-0.8 h^{-1}
Yield of ethylene	96.03% yield (1 ton of ethylene require 1.738 ton of ethanol)
Specification of ethylene product	99.95 vol% ethylene

Table 4.5 Summary of plant description for ethanol dehydration to ethylene based on Petrobras process

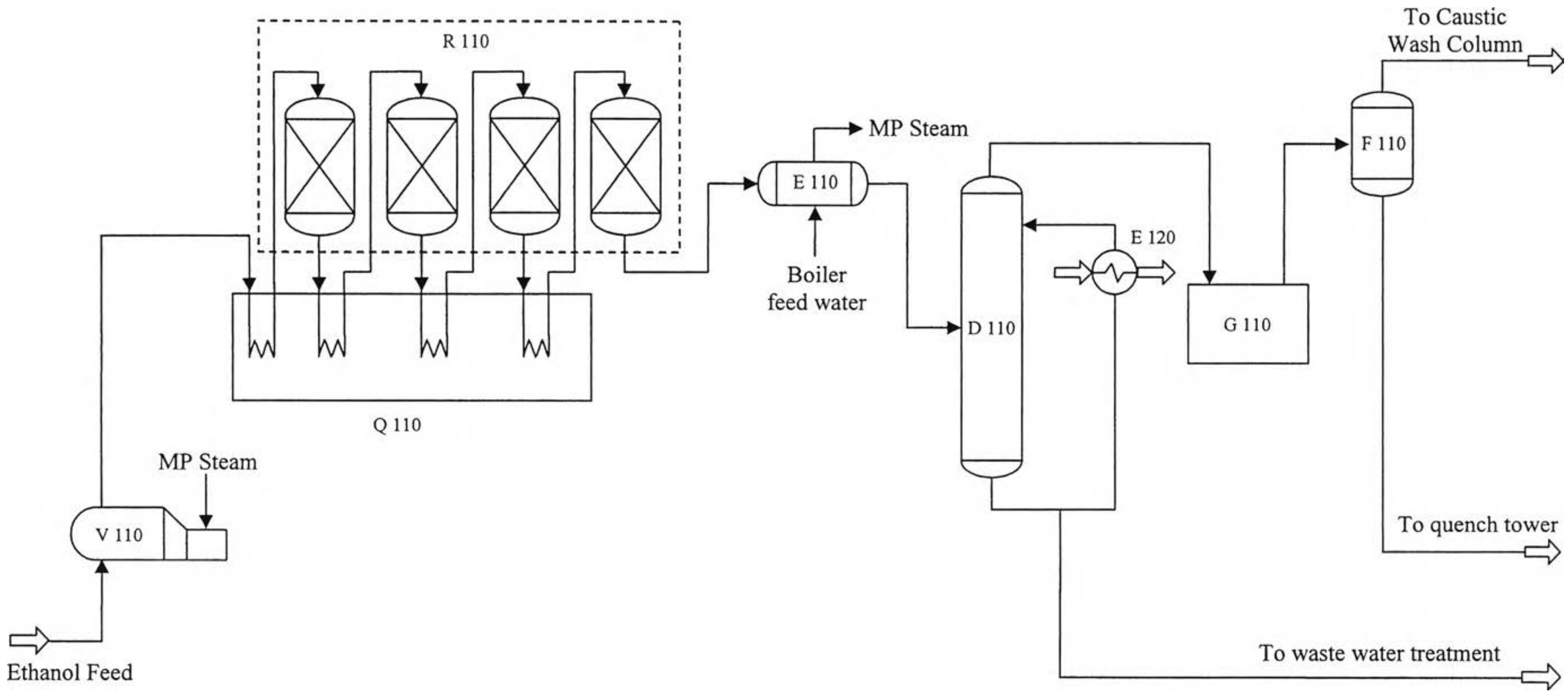
Description	Value
Plant Capacity (based on ethylene product)	200,000 tons/year 590.7 tons/day 24.6125 tons/hour
Working time (hours/year)	8,125 hours/year ; (339 days/year)
Number of adiabatic reactors	single adiabatic reactor
Catalyst	Alumina Catalyst (99% conversion and 99% ethylene selectivity)
Specification of ethanol as feed	93% ethanol (hydrous ethanol)
Operating temperature	470-485 °C
WHSV (weight hourly space velocity)	0.29-0.31 kg EtOH/hr per kg catalyst
Yield of ethylene	98.01% yield (1 ton of ethylene require 1.7698 ton of ethanol)
Specification of ethylene product	99.95 vol% ethylene

4.2.2.2 Description of Each Unit

According to the process overview of Chematur Engineering AB and Petrobras process as shown in Figure 4.3 and Figure 4.4, the plant is composed of 7 units based on the: (1) Adiabatic reaction system, (2) Quench unit, (3) Compression unit, (4) Caustic wash column, (5) Drying system, (6) Ethylene column system, and (7) Stripper, as shown in Figures 4.5, 4.6 and 4.7. Based on the function of each unit, the description of each unit can be described in three sections as follows.

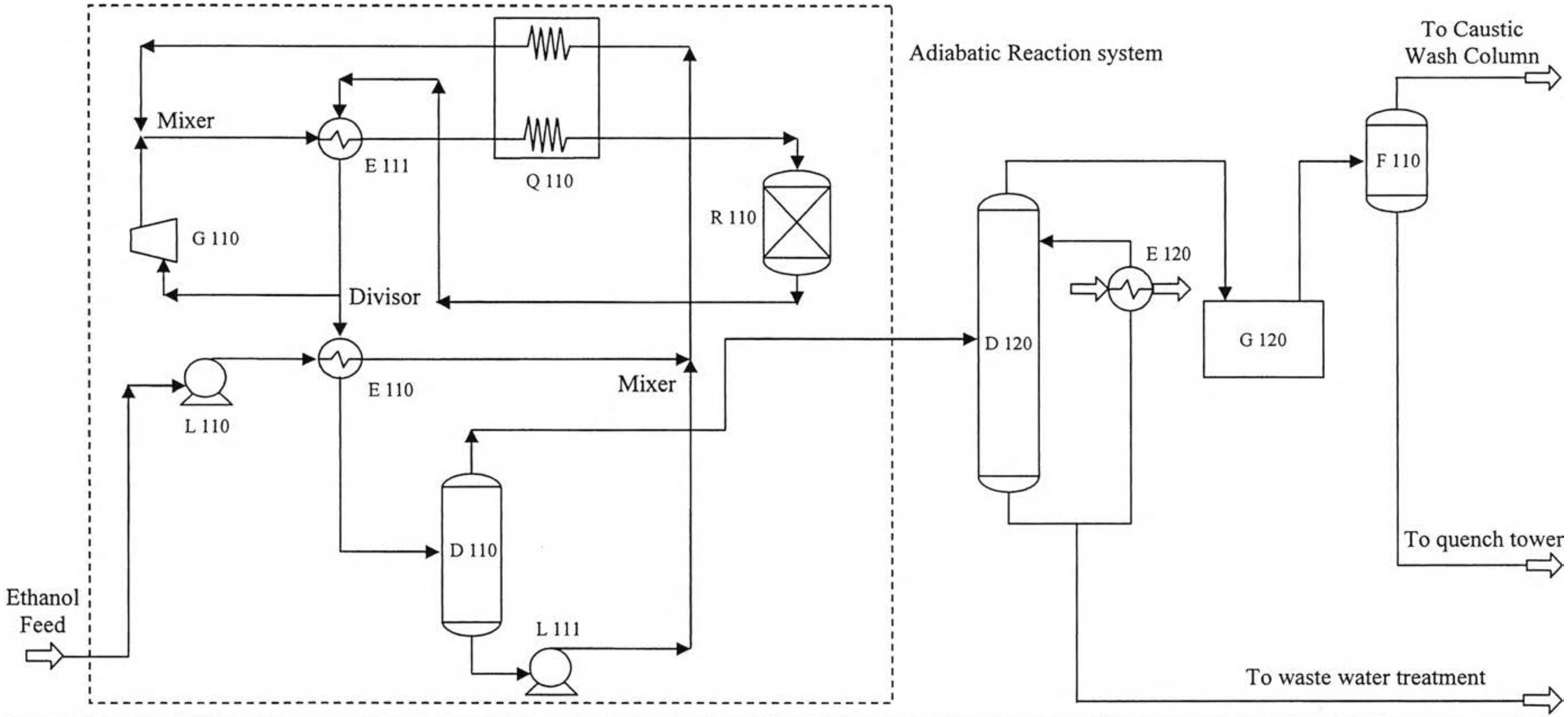
(a) Adiabatic Reaction System

Ethanol is transferred from the ethanol plant by trucks. Before processing, the ethanol is transferred to a storage tank located on-site. For the operation of **the Chematur process**, the ethanol feed from storage tank is passed to an ethanol vaporizer, V-110. The vaporized ethanol is mixed with medium pressure



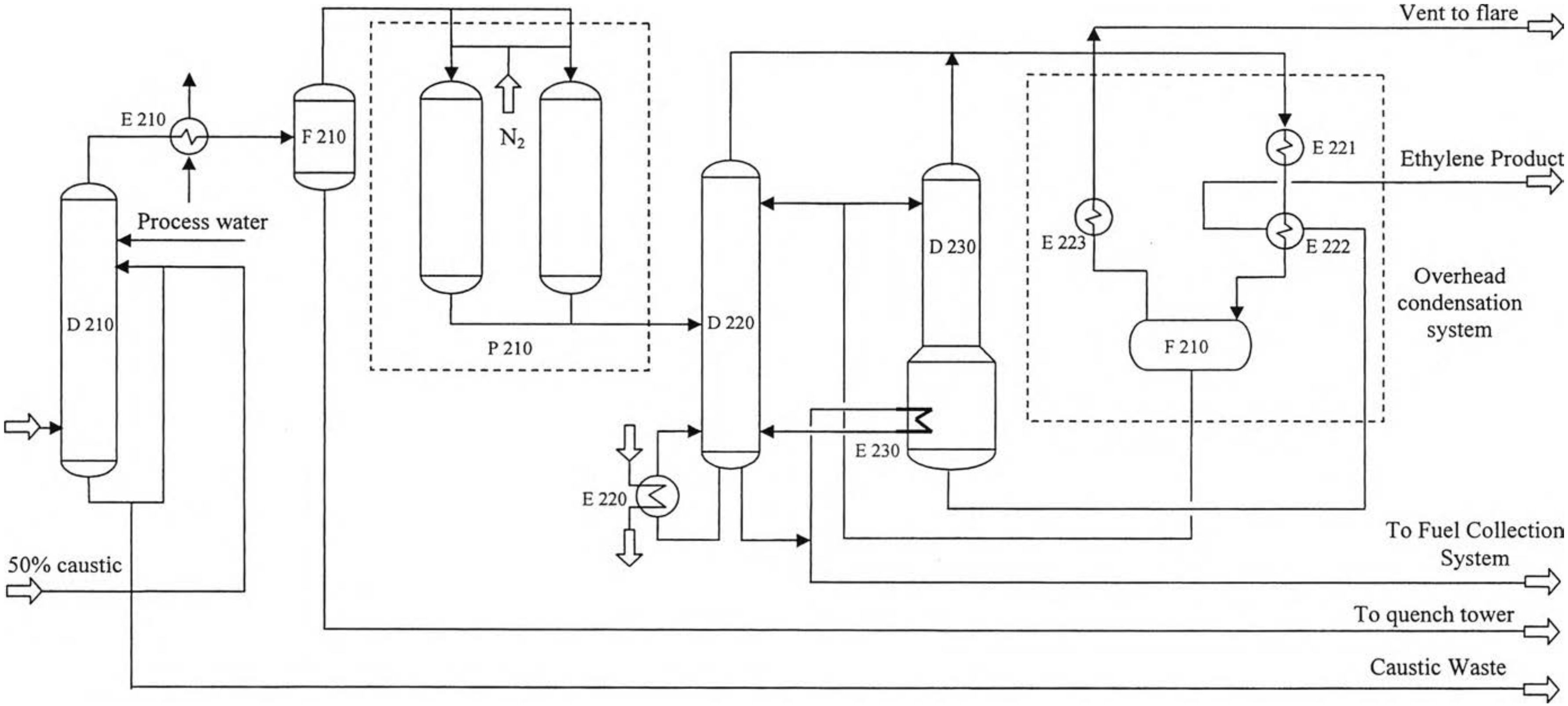
V 110	Q 110	R 110	E 110	D 110	E 120	G 110	F 110
Vaporizer	Furnace	Adiabatic Reactor System	Waste Heat Boiler	Quench Tower	Quench Cooler	Crude Ethylene Compressor	Crude Ethylene K.O Drum

Figure 4.5 A process flow diagram for adiabatic reactor system based on Chematur, quench tower and compressor.



G 110	L 110	E 110	E 111	D 110	L 111	Q 110	R 110	D 120	E 120	G 120	F 110
Recycle Compressor	Ethanol Feed Pump	Feed Heat Exchanger	Ethylene Heat Exchanger	Separating Vessel	Recycle Pump	Furnace	Adiabatic Reactor	Quench Tower	Quench Cooler	Crude Ethylene Compressor	Crude Ethylene K.O Drum

Figure 4.6 A process flow diagram for adiabatic reactor system based on Petrobras, quench tower and compressor.



D 210	E 210	F 210	P 210	E 220	D 220	E 230	D 230	E 223	F 210	E 221	E 222
Caustic Wash Column	Ethylene Dryer Feed Cooler	Ethylene Feed Dryer K.O Drum	Ethylene Dryer System	Ethylene Column Reboiler	Ethylene Column	Stripper Reboiler	Stripper	Vent Condenser	Ethylene Column Reflux Drum	Ethylene Column Condenser	Ethylene Column After Condenser

Figure 4.7 A process flow diagram for drying and purification systems of Chematur and Petrobras processes.

steam (10.4 bar.), heated in a furnace, Q-110, and passed to the first coil where its temperature is increased to a desired temperature (i.e. 450 °C). After that, the superheated ethanol then is sent to the first of four reactors in the adiabatic reactor system, R-110, filled with the **SYNDOL** catalyst bed in each reactor. In the first catalyst bed reactor, about 30% of the ethanol is converted. Because of the endothermic reaction, the effluent gas from the first catalyst bed reactor has to be reheated in the furnace, Q-110, where temperature is raised to 450 °C again, and sent to the second catalyst bed reactor. In the second catalyst bed, the ethanol conversion increases to 60% with another temperature drop. The similar fashion of reheating and reaction in the third and fourth catalyst bed increases the ethanol conversion to 86 % and finally 99%, respectively (Koh, 1993). The effluent stream leaving the reactor containing mainly ethylene (96.8%) as well as, ethane (0.5%), acetaldehyde (0.2%), propylene (0.06%), butylenes (2.4%), etc. (0.04%), is then passed to a waste heat boiler, E-110, where the heat is recovered from the hot gases for generating medium pressure steam.

For the operation of **the Petrobras process**, the ethanol feed from storage tank is pumped to the reaction system at a pressure around 13.73 bar, then vaporized in a heat exchanger, E-110, by thermal exchange with a part of the reactor effluent (ethylene and water vapor) at a temperature of 178 °C, and then mixed with the flow of recycle water from the separation vessel, D-110, where water is condensed and separated. The mixture of ethanol and recycle water is heated in the first section of a furnace, Q-110, and then fed to a mixer, where it is mixed with the compressed gaseous portion of the effluent recycle flow (ethylene vapor/water vapor), and then passed to an heat exchanger, E-111, where the mixture is heated by thermal exchange with total reactor effluent to a temperature of 297 °C. The mixture of ethanol, recycle water and ethylene vapor/water vapor is sent to a section of a furnace, Q-110, for heating up to reaction temperature in a range of 460 °C to 500 °C, and then it is fed to the fixed bed adiabatic reactor, R-110, filled with the alumina catalyst bed. The fixed bed adiabatic reactor operates at a pressure in a range of 9.8 to 15.7 bar and a temperature in the range of 470 °C to 485 °C (Barrocas and Lacerda, 2007).

The reactor effluent that is placed at a temperature of 397 °C is firstly cooled by a heat exchange, E-111, with the mixture flow (ethanol, recycle water and ethylene vapor/water vapor) being heated in a furnace. After being cooled, the reactor effluent is sent to the flow divisor, where it is divided into two flows: (a) a first flow, which is composed of gaseous recycle (ethylene and water vapor), is compressed in a compressor, G-110, and then mixed with the flow of ethanol and recycle water which it is directly fed to the reactor, (b) a second flow is passed through the heat exchanger, E-110, where ethanol is preheated, and then fed to the separating vessel (Barrocas and Lacerda, 2007). At the separating vessel, D-110, the flow of crude ethylene leaving at the top of vessel is sent to the quench tower, D-120, whereas the aqueous recycle flow (water) at the bottom of vessel is compressed in a pump, L-111, and then mixed with the ethanol flow in the mixer.

(b) Quench and Compressor

The cool gas is sent into a quench tower, D-110, where it is cooled to about 43 °C with spraying water from the top of the quench tower. In the bottom of the quench tower, the liquid phase contains mainly water (99%), unconverted ethanol, and higher boiling point acidic impurities (i.e. acetic acid). A fraction of water at the bottom of tower is cooled through a quench cooler, E-120, and circulated to the quench tower, D-110, while the remaining fraction is sent to a waste water treatment unit.

The off-gas leaving the top of quench tower, D-110, is fed to a crude ethylene compressor, G-110, where it is compressed to 17.27 bar in order to enable sufficient pressure through downstream units, and then sent to a crude ethylene knock-out drum, F-110, where the condensate is removed from the gas. The condensate is then sent to the quench tower, D-110.

(c) Drying and Purification

The cool and compressed gas from the crude ethylene knock-out drum, F-110, next flows to a caustic wash column, D-210, where carbon dioxide (CO_2) is absorbed by contacting the gas with 50% sodium hydroxide (50% NaOH) in a two packed beds in order to reduce the carbon dioxide content to a polymer-grade specification level (10 ppm. or less). On the top of the caustic wash column, D-210, process water is supplied to wash the off-gas. As a result, the trace of caustic is washed out. The caustic waste at the bottom of caustic wash column, D-210, is sent to waste disposal. The gas leaving from the top of caustic wash column, D-210, flows to an ethylene dryer feed cooler, E-210, where it is cooled to 15.6 °C by exchanging with ethylene product stream, and then sent to an ethylene feed dryer knock-out drum, F-210, where the condensate from the cooling is separated, and sent to the quench tower, D-110. After that, the cool gas is sent to an ethylene dryer, P-210, where the remaining water is removed by pressure swing adsorption using 3A molecular sieve with appropriate heating, cooling and vapor-liquid separation equipment (Koh, 1993).

After the ethylene dryer system, P-210, the dried gas is fed to an ethylene column (C_2 splitter), D-220, where ethane concentration is reduced to 400 ppm in order to meet specification level in the ethylene product. The ethylene column, D-220, containing 64 valve trays and operating at a reflux ratio of 0.6, is reboiled by an ethylene column reboiler, E-220 (Koh, 1993). All components in the bottom of this column, which consist of ethanol, diethyl ether, acetaldehyde, propylene, butenes and the major part of ethane, are sent to a fuel collection system using as a fuel for the furnace. These bottom by-products are used as a heating source for the reboiler in a stripper, D-230, as well. In the stripper, D-230, the concentration of carbon monoxide is reduced to 5 ppm., and stripping vapors are provided by a stripper reboiler, E-220.

Overhead vapor from the ethylene column, D-220, is combined with the overhead vapor from the stripper, D-230, and flows to the overhead condensation system in order to minimize the ethylene loss.

The combined vapors are sent to an ethylene column condenser, E-221, and then to the ethylene column after a condenser, E-222. The cool liquid from the ethylene column after the condenser, E-222, flows to an ethylene column reflux drum, F-210, where the vapors of by-products are separated. The vapors from the ethylene column reflux drum, F-210, containing carbon monoxide and methane removed in the stripper, D-230, are passed to a vent condenser, E-223, before being vented to fuel or flare, while the condensed phase is circulated to the ethylene column, D-220, and the stripper, D-230. The bottom product from the stripper, D-230, is the polymer-grade ethylene having a purity of 99.95 vol%.

4.2.3 Economic Evaluation

In order to perform the economic evaluation, it is necessary to have the basis assumptions of economic evaluation for the two commercial plants of polymer-grade ethylene production via the catalytic dehydration of bio-ethanol. The basis assumptions of the economic evaluation in this study are listed as follows :

(a) Plant capacity based on ethylene is 33,000 tons per year and 200,000 tons per year for Chematur and Petrobras plants, respectively.

(b) Working time for the operation is 8,000 hours per year (about 333 days per year) and 8,125 hours per year (about 339 days per year) for Chematur and Petrobras plants, respectively.

(c) Labor and maintenance cost for the two plants in each year is fixed at 3.1% of investment.

(d) The price of 95% ethanol as raw material is 22,000 baht per ton based on Thaioil ethanol price.

(e) The price of polymer-grade ethylene is 46,800 baht per ton based on PTT group's polymer grade price, Southeast Asia price.

(f) The ethanol consumption (ton of ethanol per ton of ethylene) is 1.7380 and 1.7698 for Chematur and Petrobras plants, respectively.

(g) The utilities, which are medium pressure steam, electricity, fuel, cooling water, process water, sodium hydroxide solution and nitrogen gas, are purchased. The amount of utility requirements for two processes is shown in details in Appendix C.

(h) The total operating cost of two commercial plants is assumed to be constant in each year of all economic life.

(i) All of capital expenditure is paid at a zero year.

(j) The depreciation of two plants is 20 years, and it is constant in each year of entire economic life (economic life is 20 years).

(k) The corporate tax is fixed at 30%.

4.2.3.1 Total Capital Cost Estimation (±50%)

The total capital cost of the two commercial plants of polymer-grade ethylene production via catalytic dehydration of bio-ethanol is estimated under the equipments used in the process units described in Section *4.2.2.2*, and the scope and limitations are described in Section 4.2.1. The total capital cost estimation of Chemature plant is based on the indexing method using the known capital cost and capacity obtained from Petrobras.

The total capital cost was roughly estimated in this study using the following relationship ; **Cost** $\propto$ **Size**$^{0.6}$. As a result, the total capital cost of two commercial plants is calculated as shown in Appendix D. The summary of the total capital costs of two commercial plants is shown in Table 4.6. The total capital cost of Chematur plant is 2,914 millions baht, and that of Petrobras plant is 8,618 millions baht.

Table 4.6 Total capital costs of two commercial plants for polymer-grade ethylene production via catalytic dehydration of bio-ethanol

Plants	Capacity (based on ethylene)	Estimated Total Capital Cost (Millions baht)
Chematur	33,000 tons/year	2,914
Petrobras	200,000 tons/year	8,618

4.2.3.2 Annual Operating Cost Estimation

The annual operating cost are generally divided into two groups: 1) fixed operating costs (labor and maintenance cost), and 2) variable operating costs (raw materials and utilities). The annual operating cost of two commercial plants for polymer-grade ethylene production via catalytic dehydration of bio-ethanol is summarized in Table 4.7. The details are shown in Appendix E.

Table 4.7 Annual operating costs of two commercial plants for polymer-grade ethylene production via catalytic dehydration of bio-ethanol

Items	Chematur	Petrobras
	Million baht per year	Million baht per year
Raw material cost	1,261.8	7,797.6
Labor and maintenance cost	87.420	258.540
Utility costs	173.462	690.039
Total operating cost	1,525.6	8,694.8

The annual operating cost of polymer-grade ethylene production via the catalytic dehydration of bio-ethanol for Chematur plant is 1,525.56 millions baht per year, and the one for Petrobras plant is 8,694.8 millions baht per year.

Figure 4.8 illustrates the distribution of the annual operating cost. It shows that the cost of raw material contributes the biggest portion of total operating cost for both commercial plants (about 84.6% and 89.9% for Chematur and Petrobras plants, respectively). Therefore, it should be noted that the total operating cost of the two commercial plants tremendously depends on the raw material price.

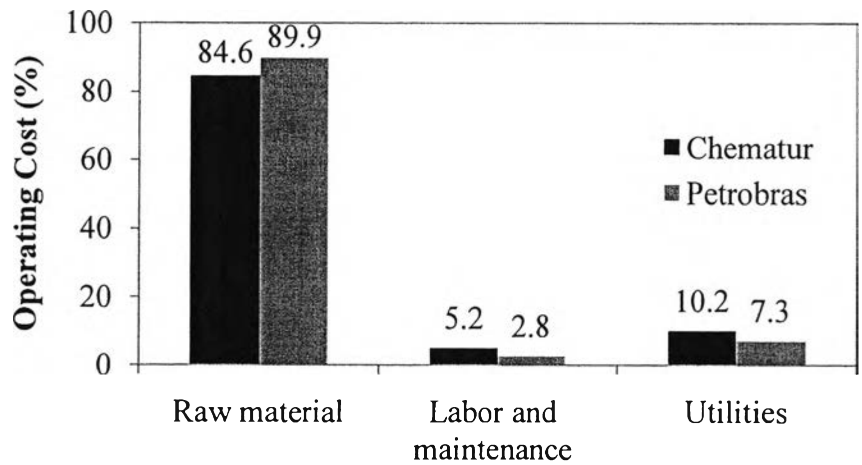


Figure 4.8 Operating cost distribution of two commercial plants for polymer-grade ethylene production via catalytic dehydration of bio-ethanol.

4.2.3.3 Annual Revenue Estimation

The revenue of two commercial plants for polymer-grade ethylene production via the catalytic dehydration of bio-ethanol only comes from the selling of polymer-grade ethylene. The selling price of polymer-grade ethylene is 46,800 baht/ton based on PTT group polymer grade price set, Southeast Asia price. The revenues of two commercial plants at two different capacities are shown in Table 4.8. The annual revenue for Chematur and Petrobras plants is 1,544.40 millions baht per year, and 9,360.00 millions baht per year, respectively.

Table 4.8 Annual revenues of two commercial plants for polymer-grade ethylene production via catalytic dehydration of bio-ethanol

Plants	Capacity (based on ethylene)	Annual revenue (Millions baht)
Chematur	33,000 tons/year	1,544.40
Petrobras	200,000 tons/year	9,360.00

4.2.3.4 Economic Evaluation

The total capital cost, annual operating cost, and annual revenue for the two commercial plants of polymer-grade ethylene production via catalytic dehydration of bio-ethanol were estimated in order to perform the economic evaluation. The total revenue minus the total operating cost gives the annual profit. The annual profits before tax and depreciation of the two commercial plants for polymer-grade ethylene production via catalytic dehydration of bio-ethanol are presented in Table 4.9. By comparing the annual profit with the total capital cost, the internal rate of return (IRR) of two commercial plants was calculated, and is shown in Table 4.9.

Table 4.9 Annual profit, internal rate of return (IRR) and payback period (PB) of two commercial plants for polymer-grade ethylene production via catalytic dehydration of bio-ethanol

Plants	Annual profit before tax and depreciation (Millions baht per year)	IRR (%)	PB (months)
Chemature	18.816	-	1,858.4
Petrobras	665.165	3.3	155.5

From Table 4.9, it can be seen that the Chematur process does not provide economic return on investment (no IRR value), which means that the Chematur process is not economically feasible, whereas Petrobras process gives 3.3% of internal rate of return. However, the project aims for 15% of internal rate of return. So, it means that the Petrobras process is not economically feasible as well.

4.2.4 <u>Sensitivity Analysis</u>

As mentioned above, the two commercial plants of polymer-grade ethylene production via catalytic dehydration of bio-ethanol are not economically feasible. Therefore, the sensitivity analysis was performed to determine the optimum parameters that make the project feasible with 15% of internal rate of return. Two parameters; that are, the cost of the raw material (ethanol price) and ethylene price, were studied in the sensitivity analysis of the two commercial plants.

4.2.4.1 Sensitivity Analysis with Respect to Ethanol Price

The analysis is performed to determine the ethanol price that makes the project feasible. The ethanol price is varied until 15% of internal rate of return is obtained, whereas the other inputs are fixed. The results of the sensitivity of internal rate of return with respect to ethanol price for two commercial plants are shown in Figure 4.9. It is observed that the decrease of ethanol price tends to increase the internal rate of return, whereas payback period (PB) decreases. The ethanol price that makes the two commercial plants feasible with 15% IRR is 9.2 baht per liter for the Chematur (57.8 months of PB) and 14.7 baht per liter for Petrobras (58 months of

PB). It can be seen that the ethanol price for Chematur plant cannot be made possible, whereas the ethanol price for Petrobras may be possible if Thaioil owns an ethanol plant. The integration of the ethanol producing plant with the ethylene producing plant could make the integrated plant commercially viable.

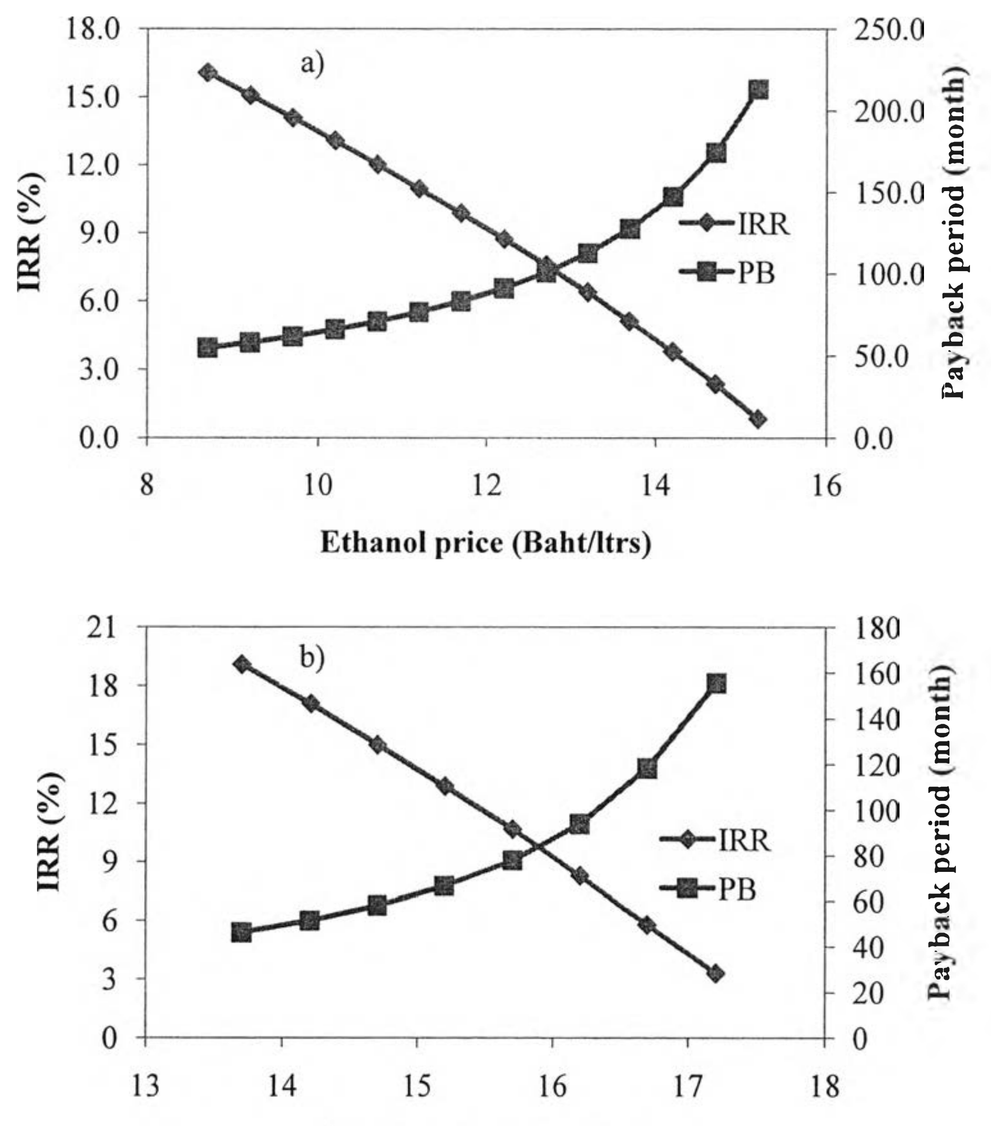


Figure 4.9 Sensitivity of IRR with respect to ethanol price for polymer-grade ethylene production via catalytic dehydration of bio-ethanol : a) Chemature plant, and b) Petrobras plant.

4.2.4.2 Sensitivity Analysis with Respect to Ethylene Price

The analysis is performed to determine the ethylene price that makes the project feasible. The ethylene price is varied until 15% of internal rate of return is obtained, whereas the other inputs are fixed. The results of the sensitivity of internal rate of return with respect to ethylene price for two commercial plants are shown in Figure 4.10.

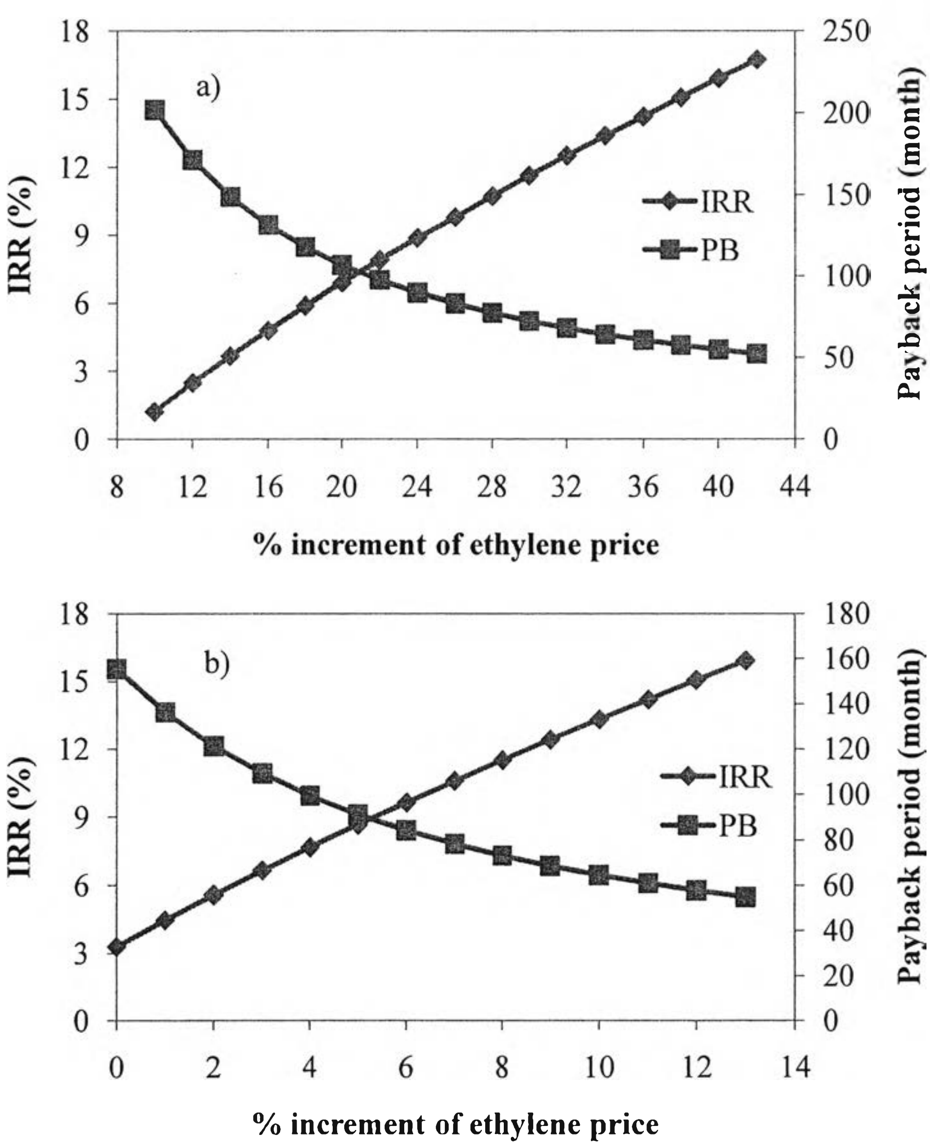


Figure 4.10 Sensitivity of IRR with respect to ethylene price for polymer-grade ethylene production via catalytic dehydration of bio-ethanol : a) Chemature plant, and b) Petrobras plant.

It is observed that the increasing of ethylene price tends to increase the internal rate of return, whereas the payback period (PB) decreases. The ethylene prices that makes the Chematur and Petrobras plants feasible with 15% IRR are 38% higher than the current price (64,584 baht per ton) and 12% higher than the current price (52,400 baht per ton), respectively.