ETHANOL TO ETHYLENE (ETE)
CASE STUDY

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A project report submitted in partial fulfilment of the requirements for the award of Bachelor of Engineering (Hons.) Chemical Engineering

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April 2012

DECLARATION

I hereby declare that this project report is based on my original work except for citations and quotations which have been duly acknowledged. I also declare that it has not been previously and concurrently submitted for any other degree or award at UTAR or other institutions.

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APPROVAL FOR SUBMISSION

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Specially dedicated to
my beloved grandmother, mother and father

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ABSTRACT

Investigation of Ethanol-to-Ethylene (ETE) process technologies for different licensors was carried out by collecting the related data such as the operating parameters, product yield, selectivity, catalyst and reactor used as well as its advantages over the conventional technologies for ethylene production. Besides, the project cost for ETE process was estimated. Simulation using Aspen HYSYS 7.1 was carried out to assess the viability of ETE process. The results showed that Lummus fluidized bed technology combined with the use of HZSM-5 catalyst is the best in term of the overall performance. It gives almost 100 % of ethylene selectivity and yield at lower temperature. Furthermore, the simulation results showed that 100 400 kg/hr of ethanol was required to produce 57 850 kg/hr of ethylene which is equivalent to 0.58 kg ethylene/kg ethanol and the energy required by the process was 2.69×10^8 kJ/hr. In addition, the total investment cost for ETE plant and the production cost of ethylene through ETE route were estimated to be USD 196 776 836 with USD 1299/ton of ethylene produced respectively. In such case, the return of investment (ROI) of 20 % and the payback period of 5 years were obtained. Lot of companies have started the production of ethylene through ETE route, especially in Brazil. However, the production cost is much higher than that of equivalent petrochemical ethylene owing to the high ethanol cost which accounts 80 % of the production cost. In conclusion, the development of cheap and sustainable conversion processes of low cost lignocellulosic biomass is crucial to enhance the viability of ETE process.

TABLE OF CONTENTS

DECLARATION **ii**
APPROVAL FOR SUBMISSION **iii**
ACKNOWLEDGEMENTS **vi**
ABSTRACT **vii**
TABLE OF CONTENTS **viii**
LIST OF TABLES **xi**
LIST OF FIGURES **xiii**
LIST OF SYMBOLS / ABBREVIATIONS **xv**
LIST OF APPENDICES **xvii**

CHAPTER

1 INTRODUCTION **1**
1.1 Background 1
1.2 Problem Statement 2
1.3 Proposed Solution 4
1.4 Aims and Objectives 4

2 LITERATURE REVIEW **6**
2.1 Cracking Technologies for Ethylene Production 6
2.1.1 Steam Cracking 6
2.1.2 Catalytic Cracking of Naphtha 7
2.1.3 Cracking of Vegetable Oils 9
2.2 Alternative Process Technologies for Ethylene Production 10
2.2.1 Methanol-to-Olefins 10

2.2.2 Ethanol to Ethylene 12

3 METHODOLOGY 20
3.1 Data Collection 20
3.2 Data Analysis 20
3.3 Cost Estimation and Economic Evaluation 21
  3.3.1 Total Capital Cost Investment 21
  3.3.2 Operating Cost 22
  3.3.3 Other Cost 23
  3.3.4 Sales Revenue 24
  3.3.5 Economic Evaluation 24
3.4 Simulation 25
3.5 Selection of Technology 26

4 RESULTS AND DISCUSSIONS 28
4.1 Foundation of Catalytic Dehydration of Bio-ethanol to Ethylene 28
4.2 Licensors 28
  4.2.1 Lummus Fluidized Bed Technology 28
  4.2.2 Halcon Scientific/Chematur Engineering AB Fixed Bed Technology 30
  4.2.3 SINOPEC Technology 30
  4.2.4 Petrobras Technology 31
4.3 Comparison between Licensors Technologies 33
  4.3.1 Process Main Characterization 34
  4.3.2 Catalyst Characterization for ETE Process 37
4.4 Selection of Technology and Catalyst 40
  4.4.1 Process Description 41
  4.4.2 Catalytic Dehydration of Ethanol 41
4.5 Viability of ETE process 44
  4.5.1 Availability of Ethanol 44
  4.5.2 Economic Evaluation 47
  4.5.3 Simulation 51

4.6 Comparison between ETE and Naphtha Cracking 56
4.7 Commercial ETE Plants and Research Activities 60

5 CONCLUSION AND RECOMMENDATIONS 62
5.1 Conclusion 62
5.2 Recommendations 63

REFERENCES 64

APPENDICES 72

LIST OF TABLES

TABLE	TITLE	PAGE
2.1	Typical Conditions for Naphtha Cracking	6
2.2	Products from Steam Cracking of Naphtha	7
2.3	Product Yields from Catalytic Cracking of Naphtha	8
2.4	Ethanol Yield for Different Crops	14
3.1	Estimation of Labour-related Operating Cost and Maintenance Cost	23
3.2	Estimation of Fixed Cost	24
4.1	Operating Condition and Product Composition of Lummus Technology	29
4.2	Comparison between Licensors Technologies	36
4.3	Typical Composition of Polymer Grade Ethylene	44
4.4	Estimation of Total Capital Investment	47
4.5	Estimation of Production Cost	48
4.6	Economic Analysis Review	51
4.7	Material Balance by Stream	53
4.8	Material Balance by Component	53
4.9	Energy Balance by Stream	53
4.10	Summary of Simulation Results	54
4.11	Comparison between Naphtha Cracking and ETE	59

A.1 Ethanol Production in Thailand 73

A.2 Ethanol Production in Indonesia 74

A.3 Ethanol Production in Philippine 75

A.4 Ethanol Production in Vietnam 76

A.5 Ethanol Production in Myanmar 76

LIST OF FIGURES

FIGURE	TITLE	PAGE
1.1	Global Oil Supply and Demand, Hike in Oil Prices	2
1.2	Global Demand of Commodity Polymer	3
2.1	Flow Diagram of an Ethylene Plant Using Liquid Feeds	6
2.2	Products Yield at Different Feedstocks	9
2.3	Process Flow Scheme of MTO Process	11
2.4	Process Flow Diagram of UOP/HYDRO's MTO Process	11
2.5	Flow Chart with the Main Raw Material and Process Used for Ethanol Production	14
2.6	Typical Flow Diagram of ETE Process	15
2.7	Ethanol Conversion over Temperature for Different Catalysts	16
2.8	Catalytic Activity of SAPO Catalyst over Liquid Space Velocity	17
2.9	Effect of Temperature and Space Time on Selectivity towards Ethylene and Diethyl Ether in Dehydration of Ethanol	18
2.10	Effect of Feed Composition on Catalyst	19
4.1	Process Flow Diagram of Lummus's Fluidized ETE ProcessF	29
4.2	Process Flow Diagram of Halcon Scientic/Chematur Engineering AB's ETE Process	30

4.3 Process Flow Diagram of SINOPEC's ETO Process 31

4.4 Petrobras Adiabatic Fixed Bed Process 32

4.5 Block Diagram of Petrobras's Co-processing Process 33

4.6 Production Distribution of Different Catalysts 40

4.7 Lummus Fluidized Bed Reactor and Regenerator 42

4.8 Ethylene Column and Striper Section 43

4.9 World Ethanol Production 45

4.10 Production Capacity in South East Asia by Country (million litres) 46

4.11 Breakeven Point Chart 50

4.12 Overall Process Design Diagram for Ethylene Production from Dehydration of Ethanol 55

4.13 Total Energy Used for Different Routes of Ethylene Production 57

4.14 CO_2 Emission for Different Routes of Ethylene Production 57

LIST OF SYMBOLS / ABBREVIATIONS

C	total production cost, USD/yr
C_{alloc}	cost of allocated plants, USD
C_{DPI}	direct permanent investment, USD
C_M	module cost of equipment, USD
C_{TBM}	total bare module investment, USD
C_{TCI}	total capital investment, USD
C_{TPI}	total permanent investment, USD
D	annual depreciation, USD
F_1, F_2	factors account for site preparation, service facilities, utility facilities and related facilities
F_M	material factor
F_p	production rate of main product, lb/yr
F_{PI}	factor account for piping, instrumentation and automatic controls, and indirect costs
F_{PR}	production rate factor
i_{min}	minimum acceptable return of investment, %
S	Sales revenue, USD
t	income tax rate, %

APEC	Asia Pacific Economic Cooperation
CTO	catalyst-to-oil ratio
DW&B	direct wages and benefits, USD/yr
ETE	ethanol-to-ethylene
FCC	fluid catalytic cracking
HC	hydrocarbon
HVC	high value chemicals
IEA	International Energy Agency

LHSV	liquid hourly space velocity, hr^{-1}
M	maintenance
M&O-SW&B	maintenance and operation salaries, wages and benefits, USD/yr
MTO	methanol-to-olefin
MW&B	maintenance wages and benefits, USD/yr
NETL	National Energy Technology Lab
O	operational
PBP	payback period, year
RCC	residual catalytic cracking
ROI	return of investment, %
ROW	rest of World
SD	Scientific Design
SEC	specific energy consumption, GJ/t
VP	venture profit, USD/yr
WHSV	weight hourly space velocity, hr^{-1}

LIST OF APPENDICES

APPENDIX	TITLE	PAGE
A	Ethanol Production in South East Asia	72
B	Simulation Results	77

CHAPTER 1

INTRODUCTION

1.1 Background

Ethylene is the simplest olefin and the most important base product in manufacturing petrochemicals. It is the largest volume of organic chemical worldwide which is converted into a range of solid and liquid products with a wide range of applications. For instances, ethylene is used to manufacture ethylene derivatives and polymers such as ethylene oxide, ethylene dichloride, polyethylene and polyethylene terephthalate (PET) (ICIS, 2011). The global demand of polymers which accounts for approximately 59.3 % of total ethylene market is expected to continue to grow over the next 10 years based on Nexant Chemsystem report (2011). Besides, SRI consulting report (2011) estimated that the consumption of ethylene is continuously rising at an average rate of 3.5 % per annum.

The most common used commercial methods for producing olefins is steam cracking whereby naphtha cracking and ethane cracking represent about 45 % and 35 % of world ethylene production capacity respectively (Gielen, Bennaceur, & Tam, n.d.). This process uses the steam as the diluting agent and heat the feedstock in furnace without the presence of oxygen to break down the large hydrocarbon into the smaller ones. The major concern of this technology is coke formation during cracking process and it consumes large amount of energy which in turn, the energy cost for production has increased. Therefore, it is necessary to produce ethylene at low energy cost by adopting the alternative routes over the conventional methods.

1.2 Problem Statement

Crude oil, the raw materials of fractions that used to produce light olefins, is a non-renewable resource. However, the oil demand is increasing due to the economic growth of developing countries even though the crude oil is depleting. Based on the Oil Market Report from International Energy Agency (2009), the oil production worldwide has always been unable to meet the increasing oil demand and the oil prices are increasing as shown in Figure 1.1. In fact, oil supplies have always remained constant in terms of volume for years.

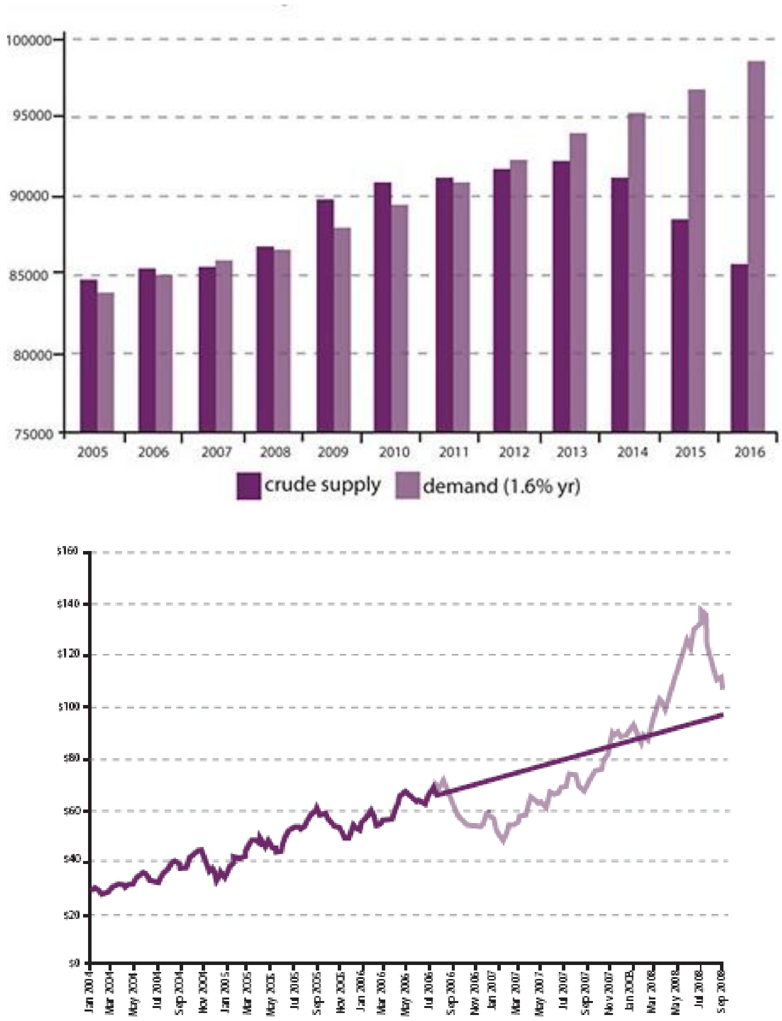

Figure 1.1: Global Oil Supply and Demand, Hike in Oil Prices

[Source: Peak Oil Consulting, 2008; IEA, 2009]

Besides, the pressure arisen from oil depletion and escalating oil demand has resulted in the hike in oil prices to reach the equilibrium in demand and supply. Currently, the oil prices are above $ 100 per barrel which has significantly burdened the investment and manufacturing sectors. As illustrated in Figures 1.2, the global commodity demand of polymers is growing and this has revived a strong interest in production of bio-based petrochemicals which is produced by using renewable feedstock to avoid the negative impact on economics and business.

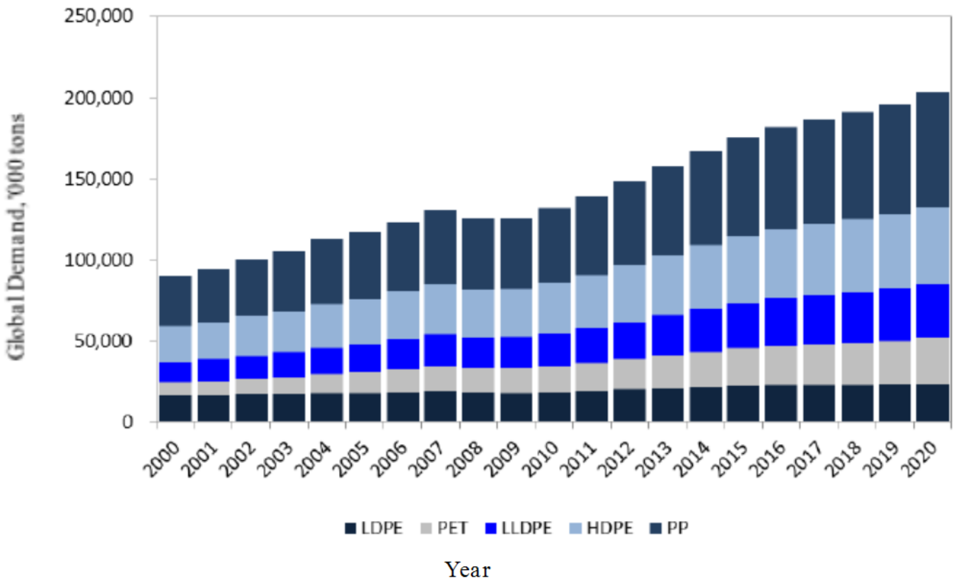

Figure 1.2: Global Demand of Commodity Polymer

[Source: Nexant, 2011]

In addition to the crude oil issues, the technologies used for ethylene production have significant impact on the environment. Steam cracking is the most energy consuming process in the chemical industry and it currently accounts for approximately 180 to 200 million tonnes of CO_2 emissions worldwide (Neelis, Patel, Blok, Haije, & Bach, 2006). Goals to develop alternative routes for ethylene production which have favourable energy consumption and greenhouse gases emission have been brought up.

In a nutshell, from both environmental and economic perspectives, it is therefore of interest to study alternative sources for ethylene production as well as energy saving potentials offered by alternative processes.

1.3 Proposed Solution

Catalytic dehydration of ethanol to ethylene so called Ethanol-to-Ethylene (ETE) was first reported since 1979 (Morschbacker, 2009). This process utilizes ethanol as feedstock which is obtained through fermentation process using agricultural products such as sugarcane, corn and biomass. Compared to the petrochemical equivalent, the main advantages of bio-ethylene are that it can reduce greenhouse gas lifetime emissions and the dependence of the chemical industry on fossil-fuels (Neelis et al., 2006).

1.4 Aims and Objectives

In this study, the alternative route to be investigated is dehydration of ethanol for ethylene production, i.e., ETE process.

The objectives of the study are as followings:

- To study and compare the licensors technologies for ETE process in term of their efficiency and propose the best technology.
- To study the feasibility of the project in term of economic worth, availability of feedstock, and practicability based on simulation results.
- To study the market distributions and market activities of the ETE process.
- To compare the ETE process with the naphtha cracking process.

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CHAPTER 2

LITERATURE REVIEW

2.1 Cracking Technologies for Ethylene Production

2.1.1 Steam Cracking

Steam cracking of hydrocarbons has been the major source of light olefins production for more than half a century. The current feedstocks for olefin production are derived from crude oil and natural gas such as naphtha and ethane. The availability of feedstock depends on the composition of crude oil and natural gas and their production volume (Rahimi & Karimzadeh, 2011). For instance, steam cracking of light naphtha produces about twice the amount of ethylene obtained from steam cracking of vacuum gas oil under nearly similar conditions.

The typical process of steam cracking and the typical operating condition are shown in Figure 2.1 and Table 2.1, respectively.

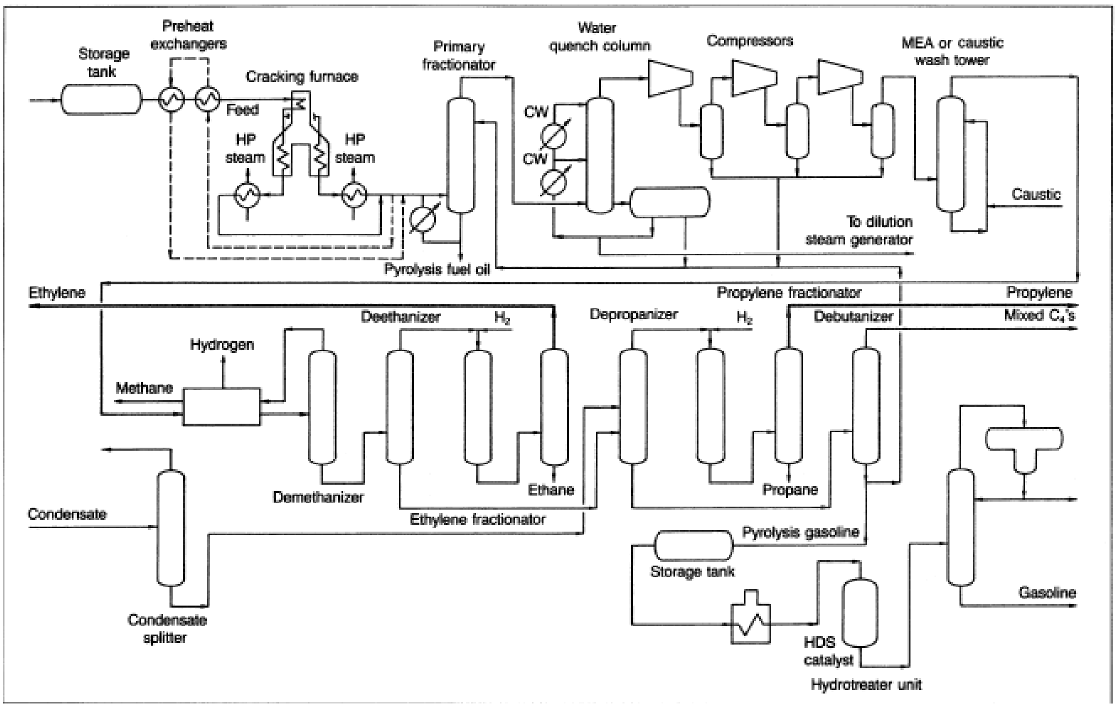

Figure 2.1: Flow Diagram of an Ethylene Plant Using Liquid Feeds

[Source: Matar & Hatch, 2000]

Table 2.1: Typical Conditions for Naphtha Cracking

[Source: Matar & Hatch, 2000]

	Condition
Temperature, °C	800
Pressure, atm.	atmospheric
Steam/HC, kg/kg	0.6-0.8
Residence time, s	0.35

Generally, liquid feed are cracked with lower residence times and higher steam dilution ratios than those used for gas feedstocks. Besides, maximum olefin yields can be obtained at lower hydrocarbon partial pressures, pressure drops, and residence times for liquid feeds (Matar & Hatch, 2000). Cracking process operated at higher severity increases the ethylene product and by-product methane and decreases propylene and butylenes. Table 2.2 shows the product distribution at low and high severity condition.

Table 2.2: Products from Steam Cracking of Naphtha
[Source: Matar & Hatch, 2000]

Products, wt %	Cracking Severity	
	Low	High
Methane	10.3	15
Ethylene	25.8	31.3
Propylene	16.0	12.1
Butadience	4.5	4.2
Butenes	7.9	2.8
BTX	10	13
C_5^+	17	9
Fuel Oil	3	6
Other	5.5	6.6

The largest energy component is the heat used in cracking which is necessary to provide the heat of reaction and the sensible or latent heat to bring the reactants to the desired reaction temperature of 750 ℃ to 900 ℃ (Gielen et al, n.d.). It accounts for 40 % of the total energy consuming every year in the entire petrochemical industry and results in high amount of CO_2 emission (Rahimi & Karimzadeh, 2011).

In addition to high energy consumption and CO_2 emission, a typical steam cracking process which cracks the naphtha into smaller molecules in gaseous state has a problem of coking that will cause inefficient of steam cracker (Tao, Patel, & Blok, 2006). Great efforts have been dedicated to the researches on developing a novel process that can overcome the deficiencies of steam cracking.

2.1.2 Catalytic Cracking of Naphtha

The current method of producing olefins via steam cracking of naphtha has several drawbacks such as the high energy consumption, the deposition of coke in the tubes, and the relatively low selectivity in ethylene from heavy feeds. This leads to catalytic pyrolysis has been studied extensively to overcome the shortcoming of steam cracking.

Catalytic cracking is a process in which the heavy hydrocarbon molecules are converted into lighter molecules by contacting the heavy hydrocarbon with the zeolite catalyst. From the Table 2.3, it was noted that the ethylene yield is lower while the propylene yield is higher in comparison with conventional steam cracking process (Nexant, 1997).

Table 2.3: Product Yields from Catalytic Cracking of Naphtha

[Source: Nexant, 1997]

Patent Number	**6346062**	**6192136**	**6192135**	**6192134**	**6199707**
Feed conversion	96.5	95.8	93.6	97.1	95.9
Ethylene, wt %	22	22.8	21.9	22.3	21.8
Propylene, wt %	22.2	24.5	23	20.8	22.4
C_4 – C_6 aromatics, wt %	26.4	20.3	20.2	22.8	21.3
Total, wt %	70.6	67.6	65.1	65.9	65.5

The performance of fluid catalytic cracking (FCC) unit is dependent on a large number of parameters which influence the conversion process in their own way. The parameters studied include the feed composition, residence time, temperature, catalyst-to-oil ratio (CTO), hydrocarbon partial pressure, catalyst properties, and riser hydrodynamics in order to find out the optimal condition for conversion process (Dupain, Makkee, & Moulijn, 2006).

There are several studies on the optimization of FCC processes. Lid and Strand (1997) reported the implementation of on-line optimization and model-predictive control to a residual catalytic cracking (RCC) unit. While Ellis, Li & Riggs (1998) presented an optimization model by combining an empirical yield prediction model for cracking products with macroscopic mass and energy balances for the unit.

2.1.3 Cracking of Vegetable Oils

Researches on various alternative sources as crude oil substitute for ethylene production were carried out to avoid the problems associated with crude oil. Zamostny, Belohlav, & Smidrkai (2011) have disclosed the use of vegetable oils which is one of the premium renewable resources as crude oil for short alkenes production via steam cracking. Figure 2.2 shows the product yield obtained based on different feedstocks used in steam cracking process.

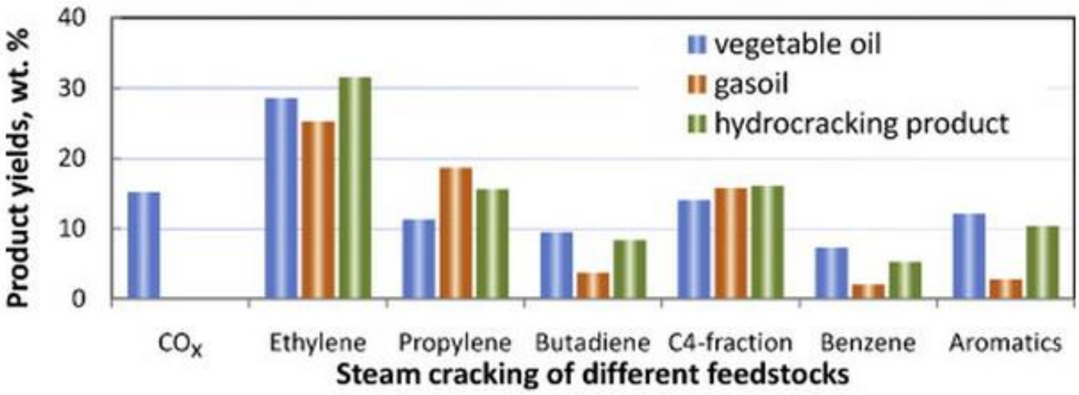

Figure 2.2: Products Yield at Different Feedstocks

[Source: Zamostny et al., 2011]

Vegetable oils form the similar products as traditional crude-oil-based feedstocks under the condition matching with the gas oil steam cracking. Besides, the long linear chain of vegetable oil leads to comparatively high content of ethylene as well as propylene and butadiene in a product mixture. Since the mechanisms of pyrolysis reactions of both the rapeseed oil and crude oil fractions are similar, it would be possible to employ well-known hydrocarbon pyrolysis technologies to vegetable oil cracking (Zamostny et al., 2011).

In addition, multiple studies regarding the topic have been published. For instance, Bielansky et al. (2010) and Dupain, Costa, Schaverien, Makkee, & Moulijn (2007) have studied on catalytic cracking of rapeseed oil to high octane gasoline and light olefins. The researches again found that the vegetable oils can be used in fuel production and petrochemical industries and form similar products as traditional crude oil feedstock under the comparable operating condition. Furthermore, the vegetable oil can be blended with the feed of an existing FCC unit at which no synergetic effects between both feeds appear to be present.

As a result, catalytic cracking of vegetable oil mixed with conventional FCC feedstock is a potential alternative for increasing the share of renewable fuels in the market. However, this alternative should be evaluated further in terms of technical, economic viability and environmental impacts since the products contain large amounts of aromatics (Dupain et al., 2007).

2.2 Alternative Process Technologies for Ethylene Production

2.2.1 Methanol-to-Olefins

Steam cracking and fluid catalytic cracking are unable to cover the expected demand for propylene. Thus, new technologies such as methanol-to-olefins (MTO) are developed due to the fact that MTO can provide a wider and more flexible range of ethylene to propylene ratio to meet market demand. The process of MTO has shown in Figure 2.3 at which the methanol is mainly produced from synthesis gas which can be obtained from any source of carbon-containing materials such as natural gas, coal or biomass (George, Goeppert, & Prakash, 2009). In addition, MTO can provide much lower costs of production and higher returns on investment than naphtha crackers and ethane crackers.

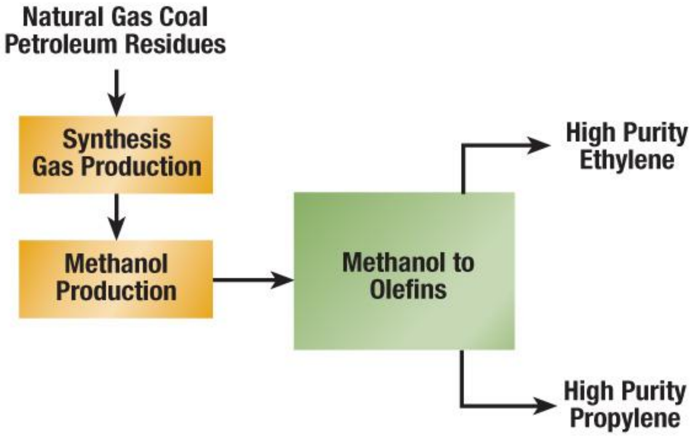

Figure 2.3: Process Flow Scheme of MTO Process

[Source: UOP, 2011]

The MTO process employs a turbulent fluid-bed reactor system at which the coked catalyst is continuously withdrawn and burned in a regenerator to burn off the coke formed. It gives 99 % conversion of methanol and having lower magnitude of coke yield and catalyst circulation than in FCC (Othmer, 1999). Figure 2.4 shows a simplified flow diagram of UOP/HYDRO's MTO process which is one of the licensors.

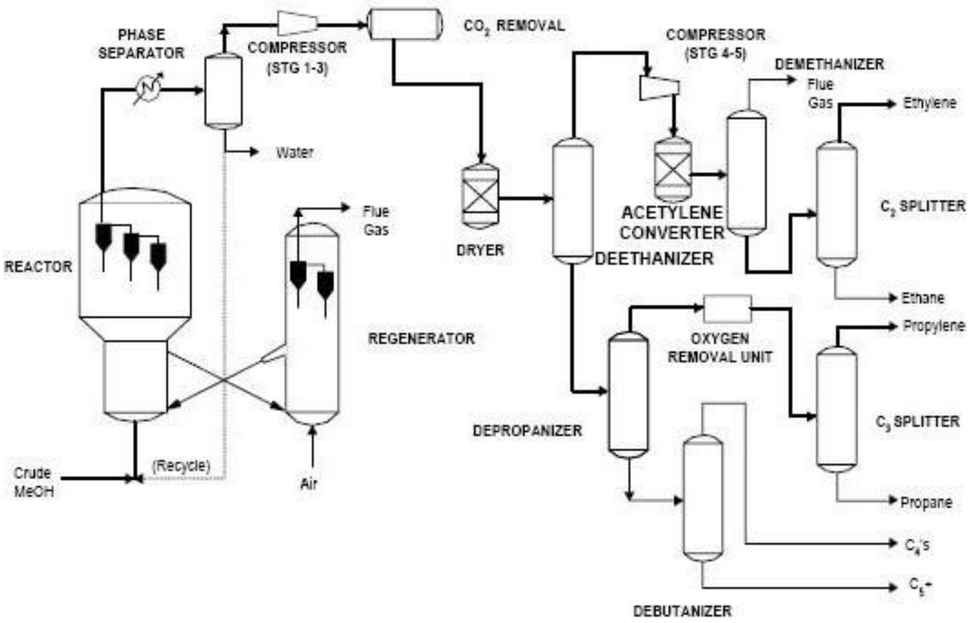

Figure 2.4: Process Flow Diagram of UOP/HYDRO's MTO Process

[Source: NETL, n.d.]

The MTO process is an acid catalyzed reaction of which SAPO-34 and ZSM-5 are the most common catalysts. However, SAPO-34 was observed to be a more efficient catalyst in terms of activity and selectivity for light olefins (Wilson & Barger, 1999). Since SAPO-34 seems to be a desirable catalyst for MTO, detailed research on the products change with the reaction conditions applying SAPO-34 catalyst and modification of SAPO-34 have been done.

For instance, Wilson & Barger (1999) have studied on characteristic of small-pore molecular sieves as catalysts for the conversion of methanol to light olefins (MTO). Kwong, Ho, Soon & Gon (2009) have investigated the effect of crystallite size of SAPO-34 catalysts on their catalytic performance. Dubois et al. (2002) have studied the catalytic performance of the modification of SAPO-34 with transition metals Co, Mn and Ni and they found out that there is significant variations in the deactivation behaviour were observed among various metal-modified catalysts.

Besides, Wu, Abraha, & Anthony (2004) have studied on the effect of reaction temperature and space velocity for MTO over SAPO-34 in bench scale fixed bed reactor. SAPO-34, seems to be the most attractive catalyst for MTO, and the MTO process has a great potential to commercialization.

2.2.2 Ethanol to Ethylene

Catalytic dehydration of ethanol to ethylene utilizes ethanol as feedstocks which are obtained from agricultural products. The agricultural products are mainly categorized into sucrose containing feedstock, starchy materials and lignocellulosic biomass Balat, Balat, & Cahida, 2008). Sucrose containing feedstocks that are currently used in commercial process includes sugarcane, sugar beet and molasses which have high sugar content; starch based feedstocks include cassava, potato, corn and wheat which are polysaccharide consisting of large number of glucose joined in a unit; while lignocellulosic feedstock such as agricultural residues which consist of complex structure can be used to produce ethanol as well.

2.2.2.1 Fundamental of Bio-ethanol Production

Generally, ethanol production is based on three steps which include obtainment of a solution of fermentable sugar, fermentation of sugars into ethanol and the last step is ethanol separation and purification to obtain a concentrated ethanol. The first step is usually different based on the choice of materials used in the production (Mussatto et al., 2010).

In the process using simple sugar, only milling process is required for the extraction of sugar for fermentation. While for the process using starch materials, saccharification process is necessary before the fermentation step at which the starch is gelatinized by cooking and submitted to enzymatic hydrolysis to form glucose monomers which is fermentable by microorganism.

For lignocellulosic feedstock, it is the most promising source for ethanol production in the future as it does not interfere with the food security. However, the process is rather complicated compare to the process that using simple sugar and starch materials for ethanol production leads to higher production cost (Cheng & Timilsina, 2011). The process requires pre-treatment to render cellulose and hemicelluloses more accessible to the process of acid or enzymatic hydrolysis in order to break down the polysaccharides into simple sugar for fermentation. The pre-treatment generally involves a mechanical step to reduce the particle size and a chemical pre-treatment to make the biomass more digestible.

The differences of ethanol production processes for different feedstocks are shown in Figure 2.5 and the conversion rate of different feedstocks to bio-ethanol is shown in Table 2.4.

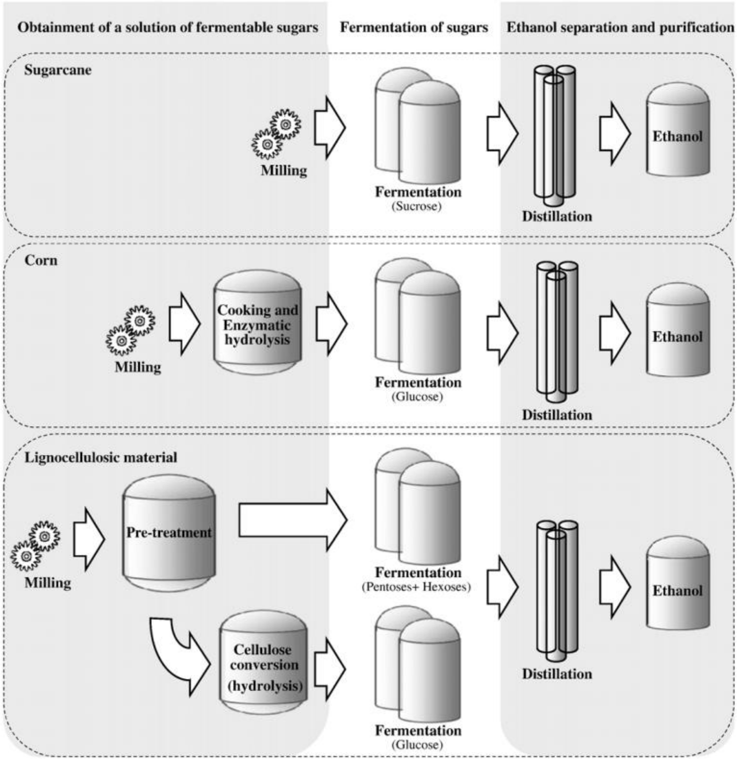

Figure 2.5: Flow Chart with the Main Raw Material and Process Used for Ethanol Production

[Source: Mussatto et al., 2010]

Table 2.4: Ethanol Yield for Different Crops

[Source: Jansson, Westerbergh, Zhang, Hu, & Sun, 2009]

Crops	Yield (ton/ha yr)	Conversion Rate (L/ton)	Bio-ethanol Yield (L/ha yr)
Sugarcane	70	70	4900
Cassava	40	150	6000
Sweet Sorghum	35	80	2800
Maize	5	390	1560
Wheat	4	450	2250

2.2.2.2 Ethanol-to-Ethylene Process

The simplified generic diagram for industrial catalytic dehydration of ethanol as showed in Figure 2.6. Depending on the catalyst characteristics, its operational age and the dehydration process used, the ethanol conversion in one reactor is usually higher than 95 %, and the reaction molar selectivity ranges from 95 % to 99 %. These parameters will have a direct influence on the raw ethylene purity and on the purification steps required to produce polymer grade ethylene.

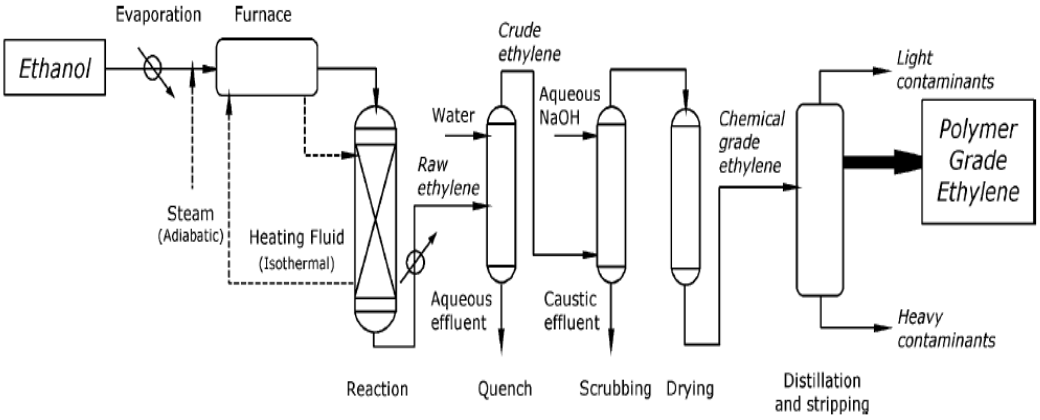

Figure 2.6: Typical Flow Diagram of ETE Process
[Source: Morschbacker, 2009]

Development of an effective catalyst is the foundational and critical step for catalytic process. Many catalysts have been studied for ethylene production via ethanol dehydration process. Doheim, Hanafy & El-Shobaky (2002) developed Na_2O-doped Mn_2O_3/Al_2O_3 catalyst which able to convert as high as 97 % of ethanol at 300 °C and at low space velocity. Chen, Li, Jiao & Yuan (2007) has also developed a catalyst doped with TiO_2 which resulted high ethanol conversion of 99.96 % and ethylene selectivity of 99.4 %.

The reactors used in industry are usually the tube-array fixed bed reactors with low liquid hourly space velocity ($0.3\ h^{-1} – 0.6\ h^{-1}$), low ethylene yield, and relatively high reaction temperature (350 °C – 450 °C), resulting in high energy consumption and low utilization of equipment capacity (Chen et al., 2007). Thus, investigations over zeolites have been done in order to develop a catalyst which allows reaction carried out at lower temperature with good catalytic performance. HZSM-5 zeolite offered an ethanol conversion of 98 % at lower temperature (300 ℃) (Phillips & Datta, 1997). Zhang, Wang, Yang, & Zhang (2008) have compared the catalytic performance of alumina catalyst, HZSM-5, SAPO-34 and modified SAPO-34 catalysts. They found out that HZSM-5 catalyst gives the highest ethanol conversion at lowest temperature which has shown in Figure 2.7.

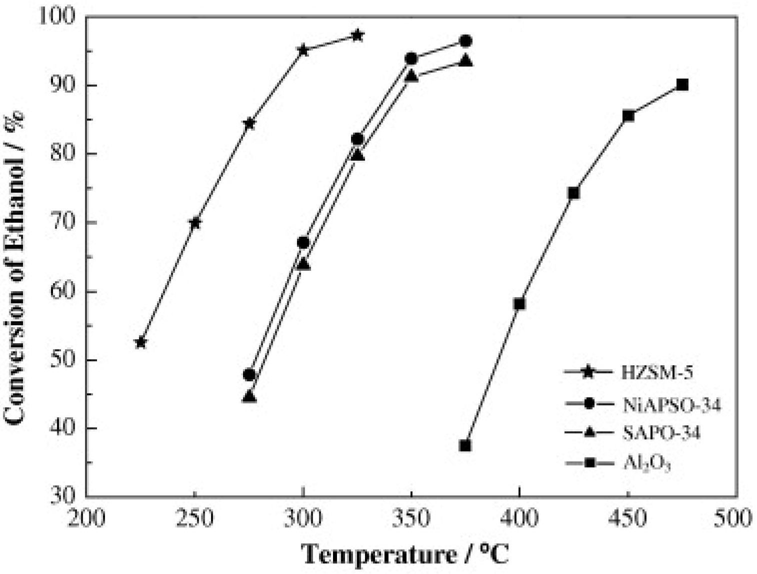

Figure 2.7: Ethanol Conversion over Temperature for Different Catalysts [Source: Zhang et al., 2008]

Effect of space time on different catalyst used was studied by Chen et al. (2010) and Chen et al. (2007). The results showed that the ethanol conversion and the ethylene selectivity decreased with rising space velocities for alumina catalyst and zeolite catalyst as shown in Figure 2.8. This is due to at the low mass space velocity, the reactant will contact with catalysts completely in the reaction, and the ethanol molecules were relatively highly activated. As the molar hourly space velocity increasing, the residence time of raw materials became shorter leads to a large number of un-reactive ethanol in products. This makes the accumulation of ethanol to form by-products easily like diethyl-ether. Therefore, the ethanol conversion and the ethylene selectivity decreased but the selectivity of diethyl-ether increased.

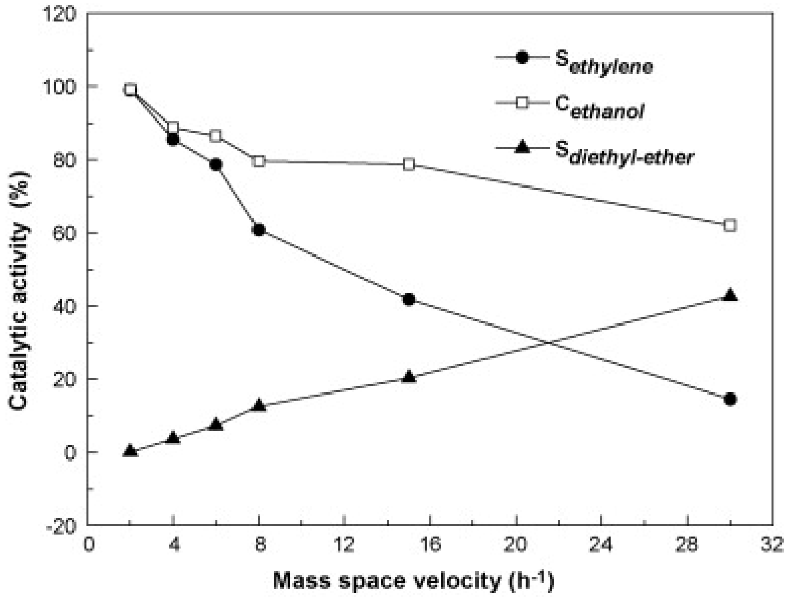

Figure 2.8: Catalytic Activity of SAPO Catalyst over Liquid Space Velocity [Source: Chen et al., 2010]

Bokade & Yadav (2011) were again showed that the conversion of ethanol increases substantially with increasing temperature at high space time. No diethyl ether was formed above 250 ℃ and the selectivity towards ethylene increased with increasing temperature and space time as shown in Figure 2.9.

In addition, presence of water in the feed stream will cause significant reduction in ethanol conversion as shown in Figure 2.10 (Chen et al., 2007). This is due to the water is adsorbed more strongly than ethanol on the catalyst surface, caused reduction in number of available active sites for the chemisorptions of ethanol for reaction. Contrary, the ethylene selectivity is increased with decreasing ethanol concentration.

However, effect of the water concentration in ethanol feed on catalytic performance was studied. Phillips & Datta (1997) stated that water in the ethanol feed enhanced the steady-state catalytic activity and ethylene selectivity by moderating the acidity of the catalytic sites, resulting in less extensive deactivation due to coking. The attenuation of deactivation by coke has been proven by measuring the coke in the MTG process on HZSM-5 zeolite (Aguayo, Benito, Ortega, Olazar, & Bilbao, 1994), in the MTO process on SAPO-34 (Aguayo, Gayubo, Ortega, Olazar, & Bilbao, 1999) and in the transformation of ethanol on HZSM-5 zeolite (Aguayo, Gayubo, Atutxa, Olazar, & Bilbao, 2002).

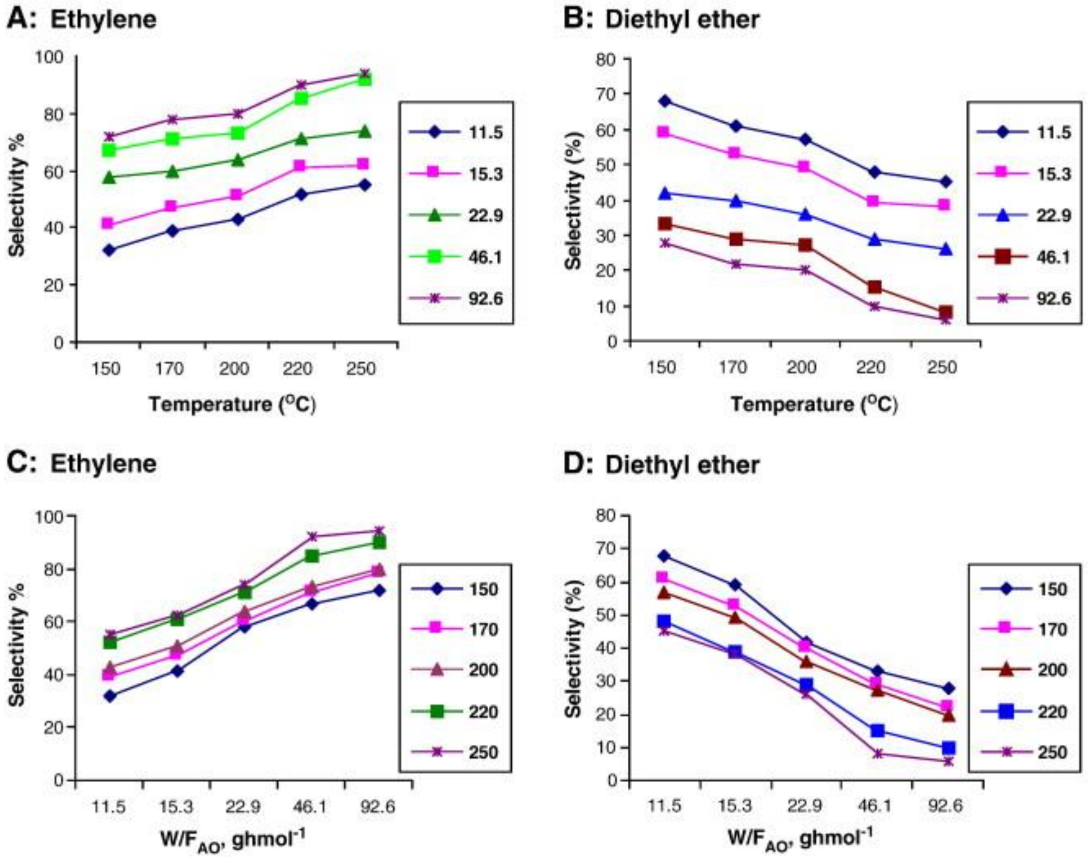

Figure 2.9: Effect of Temperature and Space Time on Selectivity towards Ethylene and Diethyl Ether in Dehydration of Ethanol [Source: Bokade & Yadav, 2011]

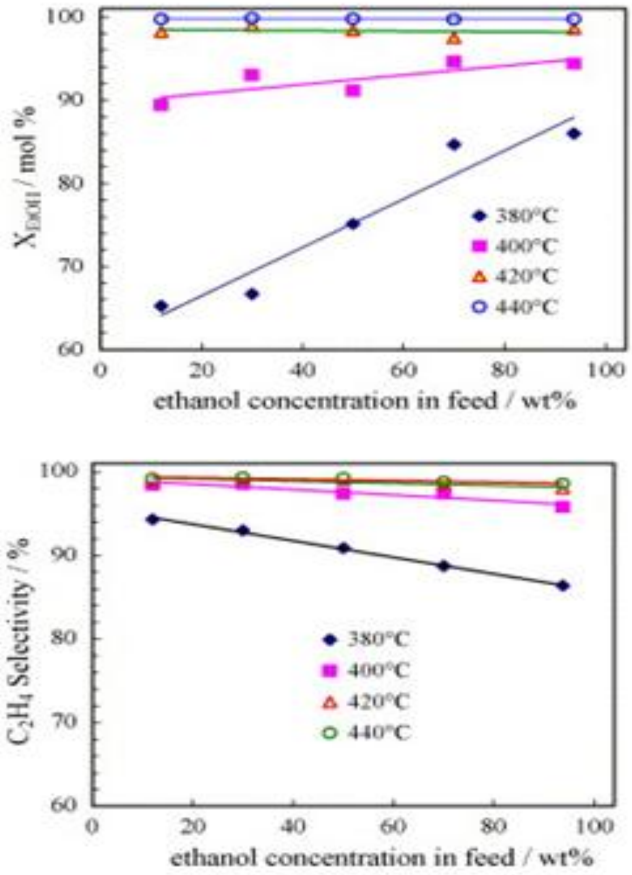

Figure 2.10: Effect of Feed Composition on Catalyst

[Source: Chen et al., 2007]

CHAPTER 3

METHODOLOGY

3.1 Data Collection

Research phase is referring to data collection phase. In this phase, library, database and internet research was used to find out all the articles and journals that are related to the project such as technology's licensors, the production process, catalyst used and the status of research. Besides, a literature case study was performed by reading all the articles that are related to the project before the project started. Through this phase, the problems are identified and further researches are required.

3.2 Data Analysis

Analysis phase is referring to evaluation of information. The data and information collected are organized into an approach to highlight the focus of the study. Besides, the performance of each technology is evaluated and the feasibility study is carried out based on the availability of feedstock, economic evaluation and simulation.

3.3 Cost Estimation and Economic Evaluation

3.3.1 Total Capital Cost Investment

An average of a low and high investment scenario and a capacity of 500 000 tonnes per year of ethylene are assumed. The Total Capital Investment is estimated using method of Hill which follows the steps below (Seider, Seader, Lewin, & Widagdo, 2010):

a. Production rate factor, F_{PR}, is computed using the sixth-ten-rule with the production rate of main product, F_p in pounds per year:

$$F_{PR} = \left(\frac{F_p}{10{,}000{,}000}\right)^{0.6} \tag{3.1}$$

b. Module cost of major pieces of equipment, C_M which include two gas compressor which assumed to operate at 200 psi; reactors which operate near atmospheric pressure; absorbers, caustic tower and two stripper which are operated atmospheric pressure is calculated using Equation 3.2:

$$C_M = F_{PR} F_M \left(\frac{\text{design pressure, psia, if} > 100\text{ psi}}{100}\right)^{0.25} \tag{3.2}$$

where

F_M is the material factor

c. Sum of the values C_M; multiply the sum by the factor F_{PI} of 2.15 to account for fluid handling piping, instrumentation and automatic controls, and indirect costs; and update with the 2010 MS cost index 1507, giving the total bare module investment, C_{TBM}:

$$C_{TBM} = F_{PI}\left(\frac{1507}{1365}\right)\sum \mathrm{C_M} \tag{3.3}$$

d. Direct permanent investment, C_{DPI} is estimated by multiplying C_{TBM} by the factors F_1 and F_2 with the values of 0.4 and 0.8 respectively to account for the site preparation, services facilities, utilities plants and related facilities:

$$C_{DPI} = (1 + F_1 + F_2)C_{TBM} \tag{3.4}$$

e. Total permanent investment, C_{TPI} and the total capital investment, C_{TCI} is estimated using the following equations, where a large contingency of 35 % is assumed and the cost of land, royalties, and plant start up are assumed to add on additional 15 %. Working capital is taken as 15 % of the total permanent investments,

$$C_{TPI} = 1.50C_{DPI} \tag{3.5}$$

$$C_{TCI} = 1.15C_{TPI} \tag{3.6}$$

3.3.2 Operating Cost

The raw material cost is estimated using the ethanol market selling price in year 2012 with the discount of 15 % due to bulk purchase. The costs for catalyst and chemicals in the ethanol dehydration production are assumed to be USD 9.2/ton ethylene produced (Arvidsson & Lundin, 2011). Other variable operating costs for the ethanol dehydration production such as utilities costs are assumed to be 2 % of total production cost.

Besides, the ethanol dehydration to ethylene plant is assumed to be operating in 4 shifts and total 6 operators per shift. They are required to work 40 hours per week and total 2080 hours per year. USD 35 per hours is estimated to be paid for each shift operator. Annual maintenance cost is estimated as a 3.5 % of total depreciable capital. The summary of the estimation of labor-related-operating cost and maintenance cost is shown in Table 3.1.

Table 3.1: Estimation of Labour-related Operating Cost and Maintenance Cost

Cost Factor	Assumption
Operation (Labor-related) (O)	
Direct wages and benefits (DW&B)	USD 35/operator
Direct salaries and benefits	15 % of DW&B
Operating supplies and services	6 % of DW&B
Technical assistance to manufacturing	USD 60,000/(operator shift) yr
Control laboratory	USD 65,000/(operator shift) yr
Maintenance (M)	
Wages and benefits (MW&B) for fluid handling	3.5 % of C_{TDC}
Salaries and benefits	25 % of MW&B
Materials and services	100 % of MW&B
Maintenance overhead	5 % of MW&B

3.3.3 Other Cost

The other cost include operating overhead, property taxes and insurance, depreciation of direct plant and allocated plant, licensing fees and general expenses. M&O-SW&B is the sum of direct wages and benefits, direct salaries and benefits of operation and maintenance. Cost of allocated plants, C_{alloc} is assumed to be 40 % of total bare module cost, C_{TBM} while the total depreciable capital, C_{TDC} is estimated to be 1.35 of $C_{DPI.}$ Table 3.2 shows the assumption used for cost estimation of ETE plant.

Table 3.2: Estimation of Fixed Cost

Cost Factor	Assumption
Operating Overhead	20 % of M&O-SW&B
Property and insurance	1.5 % of C_{TDC}
Depreciation	
Direct plant	8 % of (C_{TDC}-1.18 C_{alloc})
Allocated plant	6 % of 1.18 C_{alloc}
Rental fees	1 % of Sales
Licensing fees	2 % of Sales
General Expenses	10 % of Sales

3.3.4 Sales Revenue

Sales revenue, S was estimated by multiplying the total ethylene produced with the selling price, which is assumed to be the market price.

3.3.5 Economic Evaluation

The breakeven point is estimated by plotting a graph of sales revenue and production cost versus running capacity. The breakeven point is the intersection point where the sales revenue is equal to total production cost.

Besides, the economic evaluation is based on the return of investment (ROI) which shows the annual interest rate made by profits on the original period, payback period (PBP) which gives the time required for the annual earning to equal the original investment and venture profit (VP) which is an approximate measure of the profitability of the process. It is estimated by using equation below:

$$ROI = \frac{(1-t)(S-C)}{C_{TCI}} \tag{3.7}$$

$$PBP = \frac{C_{TDC}}{(1-t)(S-C)+D} \tag{3.8}$$

$$VP = (1-t)(S-C) - i_{min}(C_{TCI}) \tag{3.9}$$

where

t is income tax rate, %

S is total sales revenue, USD/yr

C is total production cost, USD/yr

C_{TCI} is total capital investment, USD

C_{TDC} is total depreciable capital, USD

D is annual depreciation, USD/yr

i_{min} is minimum acceptable return of investment, 20 %

3.4 Simulation

Aspen HYSYS 7.1 which provides a simulation model of catalytic dehydration of ethanol to ethylene was used to demonstrate the feasibility of ETE plant. Flash and physical property calculations for this reaction are provided by this Aspen HYSYS 7.1 process simulator. UNIQUAC thermodynamic model is used to estimates the interactions between reaction components and the process model was based on the technology selected.

3.5 Selection of Technology

The best technology is selected after the evaluation of the performance in terms of the ethanol conversion, ethylene selectivity, ethylene yield and the operating condition. Besides, the catalyst used is selected in order to get the highest yield at optimum condition.

Besides, benefits of ETE process have become more obvious by comparing with naphtha cracking after the economic evaluation in term of production cost, CO_2 emission and energy used. The term of specific energy consumption (SEC) is used to assess the energy used for the process. It is defined as energy use per ton of product which is in GJ/ton ethylene or GJ/ton HVCs. HVCs is defined as high value chemicals which include light olefin and non-olefins where by the light olefins include ethylene, propylene, and butylenes while non-olefins are referring to aromatics which produced in cracking of naphtha (Tao et al., 2008).

CHAPTER 4

RESULTS AND DISCUSSIONS

4.1 Foundation of Catalytic Dehydration of Bio-ethanol to Ethylene

Dehydration of ethanol is an endothermic reaction and requires 390 calories per gram of ethylene formed (Morschbacker, 2009). There are two competitive parallel reactions occurred during the dehydration of ethanol process:

$$C_2H_5OH \rightarrow C_2H_4 + H_2O + 44.9\ \text{kJ/mol}$$
$$2C_2H_5OH \rightarrow C_2H_5OC_2H_5 + H_2O - 25.1\ \text{kJ/mol}$$

The side reaction which produces diethyl ether is an exothermic reaction and it is favoured at low temperature, mainly between 250 ℃ to 300 ℃, while ethylene production is favoured at higher temperature, 300 ℃ to 500 ℃ (Chen et al., 2007).

With these operational conditions, acetaldehyde and hydrogen are obtained also by the reaction:

$$CH_3CH_2OH \rightarrow CH_3CHO + H_2$$

Other by-products formed by minor side reactions at higher temperature include methane, propylene, isobutanol, carbon oxide and carbon dioxide (Kagyrmanova, Chumachenko, Korotkikh, Kashkin, & Noskov, 2011)

4.2 Licensors

The ETE technologies are generally referring to catalytic dehydration of gas phase of ethanol which has been studied extensively. The reaction can be carried out in either a fixed-bed or fluid-bed reactor where by the fixed-bed route is licensed by Chematur Engineering AB/ Halcon Scientific Design and fluidized bed technology is licensed by Lummus (Nexant, 2008). There are few others ethanol dehydration process have been reported or patented which are including Petrobras Technology (Barracas, Joao, & Coutinho, 1978) and Sinopec Technology (Teng, Wang, Xie, & Gan, 2008).

4.2.1 Lummus Fluidized Bed Technology

Fluidized bed system has been developed since 1979 to improve the ethylene yield by controlling the temperature which in turn avoiding the hot and cold spots. The ethylene yield had increased to 99.5 % with ethylene selectivity of 99.6 % and ethanol conversion of 99.6 % (Tsao & Howard, 1979). The catalyst used can be any variety of dehydration catalyst such as alumina, silica-alumina, activated clays and zeolites. In this technology, silica-alumina is preferred due to its availability and it is periodically regenerated to remove the carbon and tars formed during the reaction. Lummus has dedicated a pilot plant development program for ethanol dehydration (Fariha, 2010).

The operating conditions, reactor effluent composition and process flow diagram for Lummus Fluidized Bed system are shown in Table 4.1 and Figure 4.1 based on Tsou & Howard (1979):

Table 4.1: Operating Condition and Product Composition of Lummus Technology

[Source: Tsou & Howard, 1979]

Operating Condition	
Temperature, °F	750
Average superficial velocity, ft/s	0.74
Residence time, s	2.7
Pressure, psig	9.6
Effluent Composition, mol %	
Water	50.02
Ethylene	49.75
Acetaldehyde	0.04
Ethanol	0.19

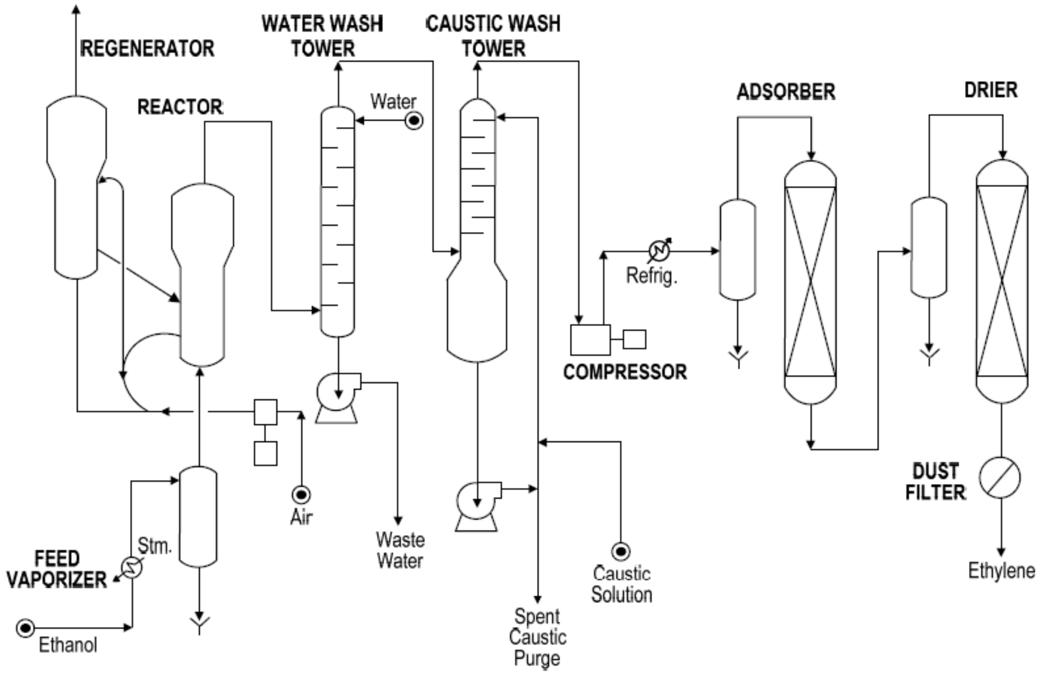

Figure 4.1: Process Flow Diagram of Lummus's Fluidized ETE ProcessF

[Source: APEC, 2009]

4.2.2 Halcon Scientific/Chematur Engineering AB Fixed Bed Technology

Chematur has developed a polymer grade ethylene production process using the fixed bed system as shown in Figure 4.2. It uses a new catalyst called Syndol catalyst which was developed by Halcon Scientific Design for their production process. Halcon claims that Syndol can stay in continuous operation for eight months without the need of regeneration and able to handle adiabatic operation (APEC, 2008). This enabled fixed bed reactor to be used and operated adiabatically at temperature range of 599 ℉ to 797 ℉. The reported ethylene yields and ethanol conversion are of 99 % and 99.8 % respectively (Fariha, 2010).

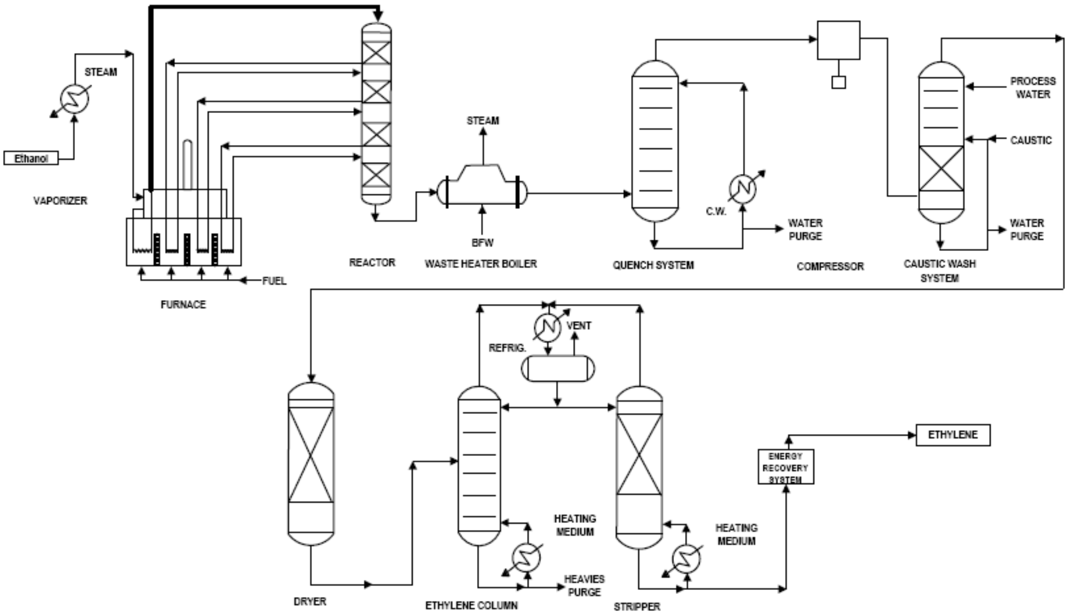

Figure 4.2: Process Flow Diagram of Halcon Scientic/Chematur Engineering AB's ETE Process [Source: APEC, 2009]

4.2.3 SINOPEC Technology

SINOPEC has developed an activated alumina catalyst which has high carbon selectivity to ethylene in fixed bed reactor. The coking tendency is suppressed efficiently and thus, no regeneration is required for this system. As a result, the ethylene selectivity and ethanol conversion obtained in this system are 97 % and 99 % in the operating temperature range of 340 ℃ to 390 ℃ according to Teng et al. (2008). The process flow diagram of SINOPEC ETE process has shown in Figure 4.3.

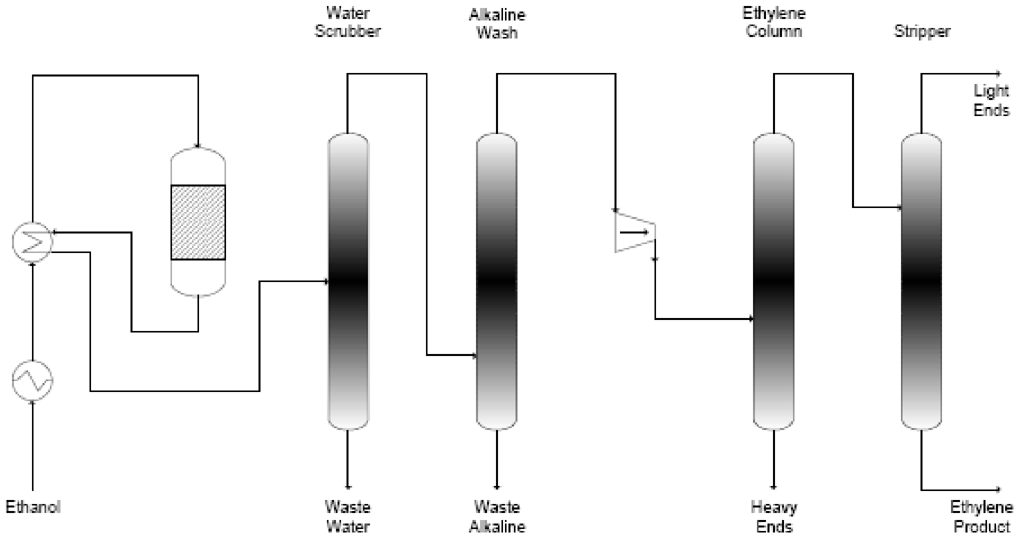

Figure 4.3: Process Flow Diagram of SINOPEC's ETO Process
[Source: Teng et al., 2008]

4.2.4 Petrobras Technology

Petrobras has developed a reactor system which is a series of fixed bed reactors that operated adiabatically. The arrangement involves simultaneous introduction of fresh feed and reactor effluent which worked as sensible heat carrying fluid between the reactors as shown in Figure 4.4 (Barracas, et al., 1978). The preferred catalyst for this invention is alumina and silica alumina due to its availability.

The adiabatic fixed-bed process had an ethanol conversion of 98 % and ethylene yield of 99.2 % at operating temperature of 355 ℃ to 390 ℃. This process increases the catalyst life at which the catalyst will sustain a period of around six to twelve months before being regenerated (Morschbacker, 2009). The process is also lower the formation of unwanted by-products, hence, is economical by saving on cost of equipments and maintenance (Fariha, 2010).

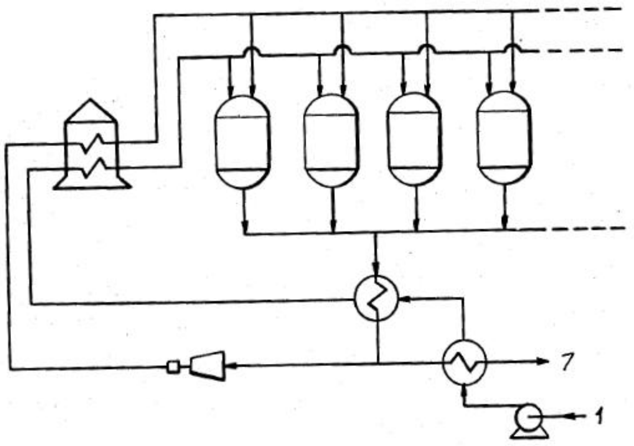

Figure 4.4: Petrobras Adiabatic Fixed Bed Process

[Source: Barracas, et al., 1978]

On the other hand, Petrobras has developed a co-processing of ethanol and hydrocarbon from petroleum refining through a fluidized catalytic cracking unit which is flexible (Rezende, Amilcar Ramos, & Fernando, 2008) as shown in Figure 4.5. Since the price and production of ethanol are varied seasonally and is dependant on the world market, it is unviable to build new units dedicated to producing ethane by the processing of ethanol alone.

In the process, the feedstocks are introduced separately into two reaction zone of the reactor and contact with zeolite catalyst for reaction to occur. Ethanol is dehydrated in a first reaction zone at temperature 530 °C to 650 °C, while the catalytic cracking of the hydrocarbons occurs in a second reaction zone at temperature of 500 °C to 620 °C. This combination gives ethylene production at 15 wt % to 90 wt % of the stream of fuel gas obtained by co-processing.

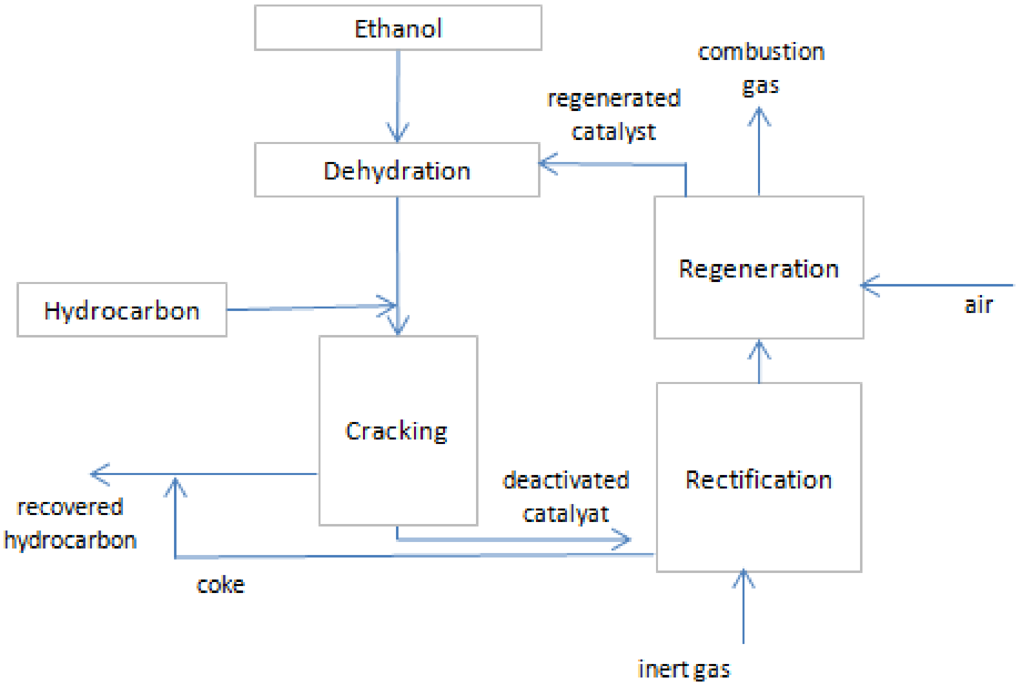

Figure 4.5 Block Diagram of Petrobras's Co-processing Process
[Source: Andrea Rezende et al., 2008]

4.3 Comparison between Licensors Technologies

The data from different technologies are tabulated in Table 4.2 in order to compare their efficiency and performance. Sinopec process and Petrobras co-processing of ethanol and hydrocarbon are new developed technologies for olefins production. They are not suitable to be compared with others technologies which focus on ethylene production. Hence, the comparison will be made between Lummus, SD/Chematur and Petrobras fixed bed reactor system.

4.3.1 Process Main Characterization

4.3.1.1 Conventional Isothermal Fixed Bed Reactor

Conventionally, industrial process for ethylene production by dehydrating ethanol is carried out using fixed bed reactor which operated under isothermal condition. The reactor is containing long parallel verticals tubes packed with the catalyst is heated externally to maintain the isotherm condition. To provide heat necessary for reaction, large number of small diameter tubes is used in order to obtain larger heat exchange area.

The operating temperatures are between 330 °C and 380 °C, with liquid hourly space velocity (LHSV) of about 0.2 hr^{-1} to 0.4 hr^{-1}. The ethanol conversion is between 98 % and 99 %, and the ethylene molar selectivity is between 95 % and 99 % (Morschbacker, 2009). This allows only working at low flow rate of vapour ethanol and leads to higher capital cost for larger throughput industrial units.

Besides, isothermal reaction control does not provide optimum selectivity or yield in all cases. As a result, such system having problems of heat transfer and it is not favour in term of economic as scale up to larger capacities is limited by the difficulty of temperature control (Rezende et al., 2008). Hence, Lummus fluidized bed system, Petrobras and Chematur fixed bed system which operated adiabatically have been developed to replace the isothermal fixed bed system.

4.3.1.2 Lummus Fluidized Bed Reactor

A fluidized bed is a situation where the solid particles are kept in suspension by a gas flow passing upwards through them. During the process, the catalyst circulates between the reactor and regenerator continuously. In the regenerator, the catalyst is restored to its initial performance by burning the coke formed on it. The heated catalyst is then introduced back into the reactor to provide the heat required by the endothermic reaction simultaneously with the hot feed. Hence, there is no external heating of the reactor.

In addition, the use of fluidized bed reactor allows continuous regeneration of catalyst permits the catalyst activity to be held constant. Fluidized system has higher specific throughput and it gives good mixing which provides temperature uniformity across the bed and avoid localized hot or cold spot. Besides, an optimum temperature control is allowed in the system at which the ethanol conversion at 400 °C is 99.5 % and 99.9 % of molar ethylene selectivity can be reached (Rezende et al., 2008). In other word, the process achieves nearly quantitative conversion of ethanol with very high selectivity to ethylene. The pilot plant demonstration was success showing that this technology is viable.

Table 4.2: Comparison between Licensors Technologies

Technology	**Lummus Fluidized Bed**	**Chematur**	**SINOPEC**	**Petrobras**	**Petrobras**
Reactor	Fluidized Bed	Fixed Bed	Fixed Bed	Fixed Bed	Fluidized
Adiabatic/Isothermal Operation	Adiabatic	Adiabatic	Adiabatic	Adiabatic	Adiabatic
Temperature (℃)	399	315 – 425	340 – 390	355 - 390	530 - 650
Catalyst	Wide variety	Syndol	Activated alumina	Wide variety	zeolite
Regeneration cycle (months)	-	8-12	12	6-12	-
Ethylene selectivity (%)	99.6	97-99	97	97-99	-
Ethylene yield (%)	99.5	> 99	-	99.2	low

4.3.1.3 Petrobras and Chematur Adiabatic Fixed Bed Reactors

Typically, this system consists of few fixed bed reactors which are in series, with additional make up between each stage and intermediate furnaces to reheat the stream that feeds the next reactor.

For Chematur process, the reaction heat is provided by using a series of catalyst beds with interheating of the reactor effluent from the first bed to higher temperature and introducing the heated effluent in a second catalyst bed and so on (Vermeiren, 2010). On the other hand, for Petrobras, it involves simultaneous introduction of fresh feed and reactor effluent which worked as sensible heat carrying fluid between the reactors (Rezende et al., 2008). Typically, the ethanol conversion is higher than 99 % and the ethylene molar selectivity is between 97 % and 99 % with an inlet temperature of about 450 °C to 500 °C.

Adiabatic fixed bed reactor requires no heating fluid circulation which will restrict the efficiency of the thermal exchange. As the consequences of steam usage, the coke build-up is smaller which leads to longer catalyst regeneration cycle. It is also allows operation to be interrupted for the exchange of catalyst or for maintenance purpose without stopping the process continuity (Rezende et al., 2008).

4.3.2 Catalyst Characterization for ETE Process

An efficient catalyst is the key technology to the industrialization of dehydration of microbial ethanol to ethylene. Reported catalysts for ethanol dehydration include silica alumina (Roca, Mourgues, & Trambouze, 1969), activated alumina (Kojima, Aida, & Asami, 1981), heteropolyacid (Ali, Al-thabaiti, Alyoubi, & Mokhtar, 2010), and zeolites (Mao, Levesque, McLaughlin, & Dao, 1987).

4.3.2.1 Activated Alumina Catalyst

Catalysts based on activated alumina are the most important kind of catalyst for bio-ethanol dehydration, and are the catalysts dominantly used in current industrial production. Scientific Design/Chematur has developed an improved alumina catalyst, Syndol which having a main composition of Al_2O_3–MgO/SiO_2. It offers high productivity that allows a reduction of total reactor volume (Chen, 2010). Besides, it has high yield conversion and selectivity which leads to better utilization of the ethanol feed and less by-products formation. The most important issue is Syndol catalyst has high resistance to deactivation and thus, it permits longer runs of operation (Chematur Engineering Group, n.d.).

Several oxide catalysts have been developed. For instance, $TiO_2/\gamma\text{-}Al_2O_3$ catalyst used for dehydration of bio-ethanol has high bio-ethanol conversion and ethylene selectivity of 99.96 % and 99.4 % respectively (Tsao et al., 2010). However, activated alumina catalyst has disadvantages of low activity, low LHSV of the feedstock, high reaction temperature, and high energy consumption (Huang, 2010). This had attracted many researchers to study on the catalytic performance of zeolite catalyst at low temperature.

4.3.2.2 HZSM-5

HZSM-5 zeolite is a high silica zeolite which consists of a regular framework with a pore size which is intermediate to the large pore sized zeolites and the small pore sized zeolites. Their shape selectivity is the catalytic expression of many factors such as the sieving effects, the geometric (shape) and size constraints on the reaction intermediates (Mao et al., 1987).

Zeolite used to replace alumina allows the employment of very dilute ethanol which can eliminate the need of costly dehydration of water content. Among the zeolite catalyst investigated, the most promising and attractive catalyst for commercial use is HZSM-5 as it could carry out the reaction at temperature below

300 ℃. the reaction temperature can be as low as 250 ℃ and the weight hourly space velocity (WHSV) ranges from 0.1 hr^{-1} to 50 hr^{-1}. The lowest temperature to achieve ethylene yields of at least 95 wt % from an aqueous ethanol (10 wt %) is 275 ℃ at WHSV of 3.2 hr^{-1} (Mao, 1989).

According to Chen et al. (2010), the increase of the weak acid sites and the acid amounts of both weak and strong acid sites, especially the amounts of weak acid sites will enhance the catalytic activity for dehydration of ethanol to ethylene, while the excessively strong acidity will lead to ethylene polymerization. In other words, reducing Bronsted acid sites and increasing lewis acid sites are advantageous for bio-ethanol dehydration to ethylene (Mao, 1989). Thus, researches have been conducted on modification of HZSM-5 such as modification with phosphorus (Lu & Liu, 2011) which controls the surface acidity active sites and surface acid strength distribution. This helps to improve the hydrothermal stability, anticoke ability, catalyst activity and selectivity.

In addition, research on nanoscale HZSM-5 catalyst below reaction temperature of 300 ℃ has been studied by Bi, Guo, Liu, & Wang (2010). They investigated the catalytic performance and thermal stability of nanoscale HZSM-5 over microscale HZSM-5 and concluded that nanoscale HZSM-5 zeolite catalyst exhibits better stability over a long period. This is attributed to the diffusion path of product over or inside the micro-catalyst is much longer than that over or inside the nano-catalyst, which leads to the carbonaceous deposits. Thus, the activity of the nanoscale HZSM-5 is passivated slowly.

4.3.2.3 SAPO-34

Another catalyst used for ethanol dehydration is SAPO-34 zeolite catalyst which is a small pore silico-alumina-phosphate consisting of an 8 membered rings system with a cage structure. Zhang et al. (2008) have compared the performance of HZSM-5 and SAPO-34 catalyst. They found out that the conversion of ethanol and ethylene selectivity for HZSM-5 is higher than that of SAPO-34 at lower reaction temperature.

However, the stability of HZSM-5 is lower than that of SAPO-34 which leads to further studies and modification of HZSM-5.

Figure 4.6 compares the product yield from different catalyst and it concluded that the ethanol conversion of Al_2O_3 is lowest and the HZSM-5 gives the highest ethylene yield amount the three catalysts.

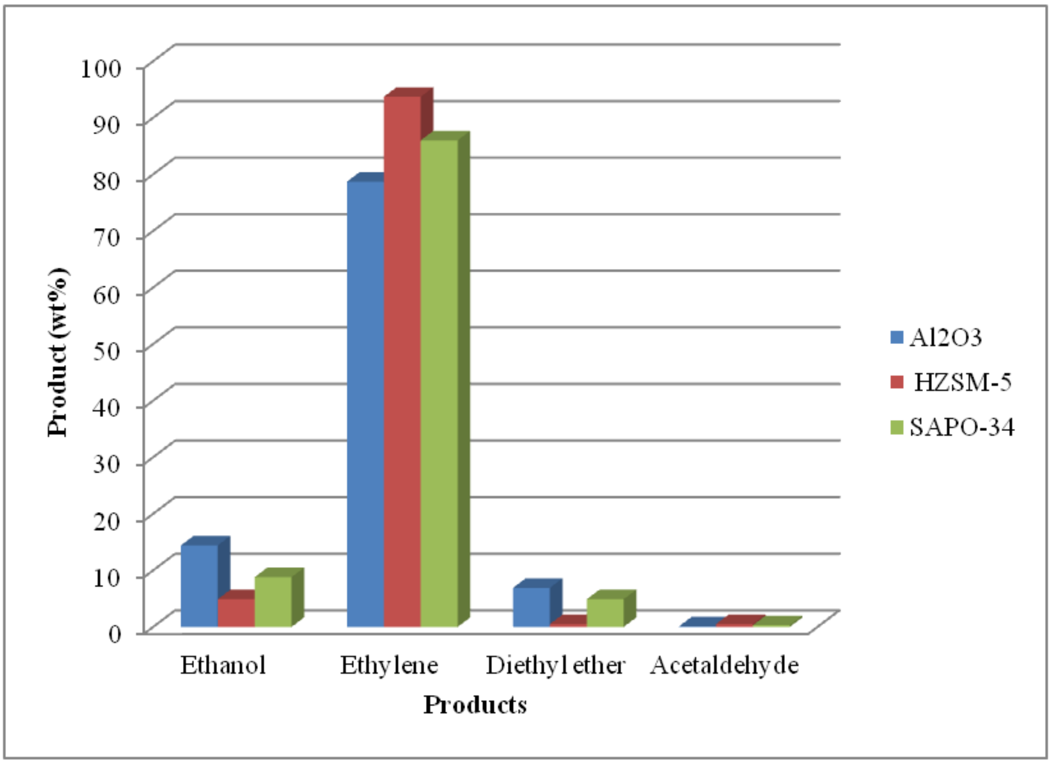

Figure 4.6: Production Distribution of Different Catalysts

[Source: Zhang et al., 2008]

4.4 Selection of Technology and Catalyst

In term of overall performance, Lummus fluidized bed technology is comparatively better as the process achieves nearly quantitative conversion of ethanol with very high selectivity to ethylene. The technology produces highest ethylene yield as fluidized bed reactor affords highly efficient temperature control, high rates of heat and mass transfer and uniform catalyst activity.

Besides, it allows continuous operation and able to process large quantities of reactants with minimum mechanical devices and operating labour (APEC, 2008). In addition, fluidized bed technology claimed that single fluidized bed reactor can produce output three times more than a single fixed bed reactor (APEC, 2008). Thus, the fluidized-bed reactor system is more economic than the fixed-bed system in a large plant.

Zeolite catalyst of HZSM-5 is preferred as its activity and selectivity are the highest among the other catalyst at the lower temperature. It gives almost 100 % of ethanol conversion and ethylene selectivity. Besides, regenerator in Lummus process which removes the coke formed on the active sites of HZSM-5 will enhance the catalyst's reactivity and stability. In addition, several studies have shown that modification of HZSM-5 with Mo species or P species had successfully improved its activity and stability. Thus, it is strongly believed that HZSM-5 catalyst is the most promising catalyst in ETE process.

4.4.1 Process Description

The process based on dehydration of vapour ethanol to ethylene consists of two steps which are dehydration step to convert ethanol to ethylene and purification step which remove water from ethylene produced.

4.4.2 Catalytic Dehydration of Ethanol

In the dehydration step, the Lummus fluidized technology is used as shown in Figure 4.7. The liquid ethanol at atmosphere pressure is pumped to the operating pressure and is vaporized to reaction temperature. The vaporized ethanol (10) is then introduced into a fluidized bed reactor (11). The reactor contains a suitable catalyst and the catalyst bed (12) is maintained in fluidized state by the gaseous ethanol fed in.

The fluidized bed reactor is maintained at dehydration conditions which are at 750 °F and 9.6 psig.

The products are withdrawn from the reactor and a part of deactivated catalyst is transported to regenerator (17) by air. The fresh catalyst in a hopper (26) is transported to regenerator together with the deactivated catalyst. The catalyst is maintained in fluidized bed state in the regenerator by the heated air which is heated by fuel in the air heater. The regenerator is operated at a temperature to heat the deactivated catalyst as to burn off the carbon and tar formed during the dehydration reaction. The heated catalyst is then introduced into the reactor to provide heat required by the endothermic dehydration reaction and the combustion by-products are withdrawn from regenerator through line 24.

To remove the traces of carbon dioxide in the gas stream, the carbon dioxide is adsorbed by washing the gas with sodium hydroxide (NaOH) in a caustic-wash tower. This is followed by compression of ethylene stream and the stream is passed through a desiccant drying bed with molecular sieves to produce chemical grade ethylene with purity higher than 99 %.

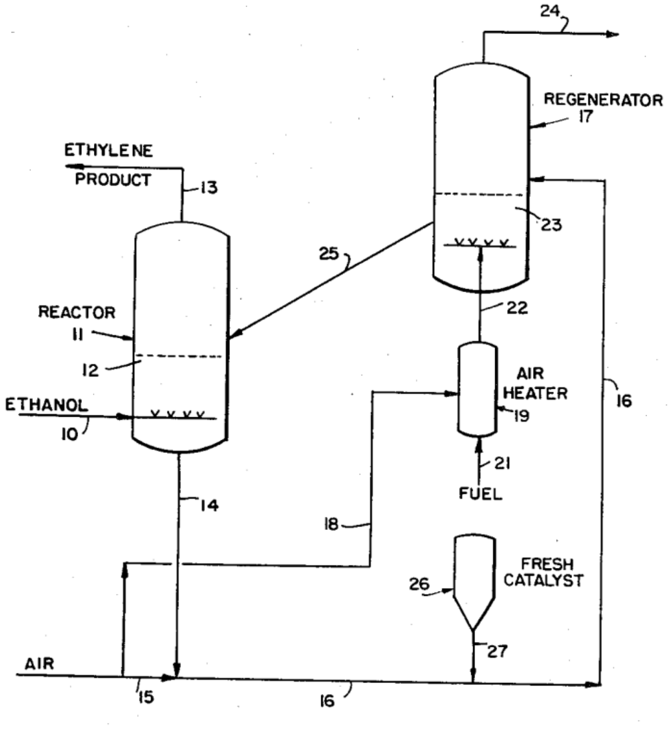

Figure 4.7: Lummus Fluidized Bed Reactor and Regenerator

[Source: Tsou & Howard, 1979]

4.4.2.1 Polymer Grade Ethylene

In order to produce polymer grade ethylene, a further treatment of ethylene stream is required as shown in Figure 4.8.

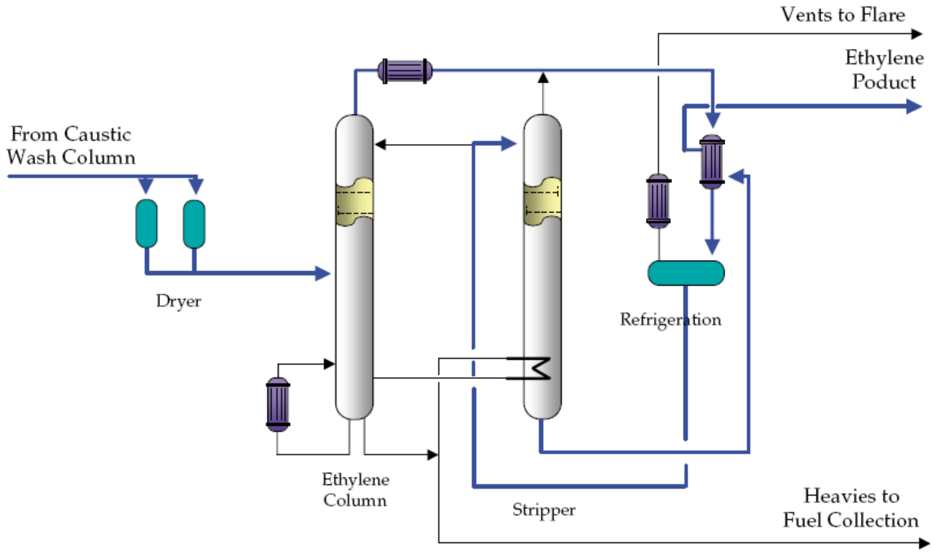

Figure 4.8: Ethylene Column and Striper Section

[Source: Chematur Engineering Group, n.d.]

The chemical grade ethylene is fractionated in a distillation and stripping systems to remove heavy and light contaminants. The bottom product of ethylene column consists of heavier carbohydrates, ethanol, diethyl ether, and acetaldehyde and the carbon monoxide, methane, and hydrogen are separated from ethylene as top products in the stripper. The light by-products separated in the stripper are vented to air while the condensed phase which is mainly ethylene is recirculated to ethylene column and stripper to obtain the final product with purity adequate for most kinds of polymerization catalytic system. Table 4.3 shows the typical composition of polymer grade ethylene.

Table 4.3: Typical Composition of Polymer Grade Ethylene

[Source: Arvidsson & Lundin, 2011]

Component	Composition
Ethylene (vol %)	99.95
Carbon Monoxide (vol ppm)	5
Carbon Dioxide (vol ppm)	10
Ethane	0.05

4.5 Viability of ETE process

A feasibility of a project is mainly focused on the aspects of economic, social and technical. The availability of ethanol was reviewed as well to ensure stable supply of ethanol for ethylene production. The economical values were covered by estimating the production cost and payback period of the investment while the technical aspects were reviewed by carrying out the simulation of ETE process to demonstrate the process viability. On the other hand, the ETE process was compared with the naphtha cracking process in order to assess the advantages of ETE process over the conventional technologies.

4.5.1 Availability of Ethanol

One of the factors that evolve the bio-ethylene production is availability of raw material. Developing countries envisage bio-ethanol as biofuel which is a potential means to improve access to energy, increase income and employment rate, alleviate poverty, spur rural development, reduce oil imports and enhance exports of biomass products. These interests render production of bio-ethanol to be growing continuously as shown in Figure 4.9 with the total production of 25 billions gallons in 2012 due to high demand of biofuels in transportation sector (Global Biofuel Center, 2012).

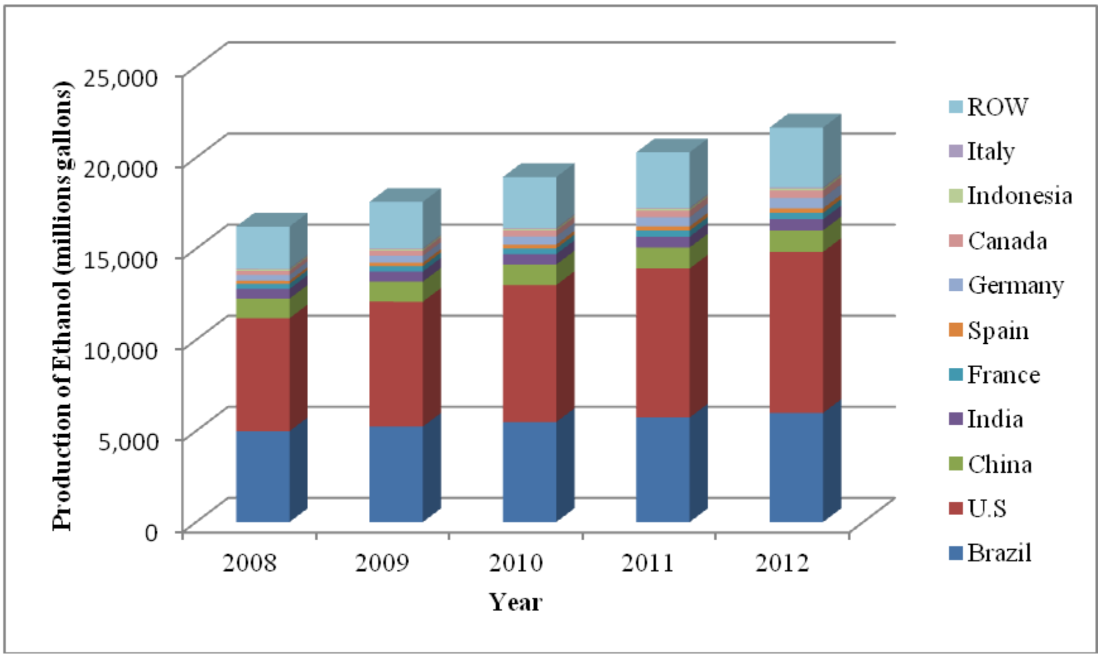

Figure 4.9: World Ethanol Production

[Source: Market Research Analyst, 2012]

On the other hand, in South East Asia, the ethanol production was estimated to reach around 7 million litres per year as illustrated in Figure 4.10 which includes the current production and the future potential production. The production of ethanol is increasing due to the encouraging national government policy, continuous government support, low production cost and simple conversion process using the sugarcane, molasses and cassava.

For instance, Thailand has successfully promoted the gasohol E10 policy leads to massive production of ethanol. This in turn leads to the start of higher blends such as E20 and E85 which have been launched in the market in recent year. Thus, the Government of Thai aimed to raise the ethanol production to three millions litres per year by 2012 which enhances the ethanol industry (Zhou & Thomson, 2009).

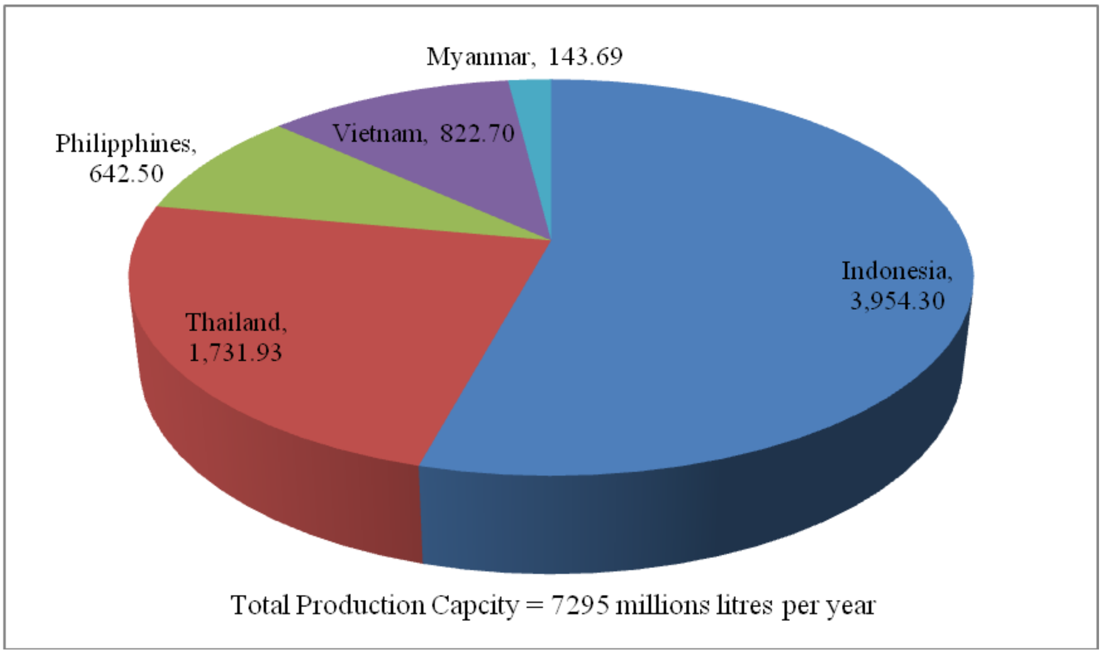

Figure 4.10: Production Capacity in South East Asia by Country (million litres)

Besides, China and Korea has started to invest on ethanol production in South East Asia. The China group developer, Henan Tianguan Group has planned to build 50 000 hectares cassava project for ethanol production with the capacity up to 86 million gallons per year in Laos (BiofuelsDigest, 2011). Conversely, MH Bio energy Group of South Korea has started the first cassava based ethanol plant in Cambodia due to its huge unused land which favors the plantation of cassava (The Bioenergy Site, 2009).

As a result, countries in South East Asia have the potential of ethylene production due to the substantial production of ethanol. Moreover, the improving technologies for ethanol production using lignocellulosic feedstocks have contributed to a lower production cost and higher ethanol productivity which in turn, enhances the bio-ethylene production as well.

4.5.2 Economic Evaluation

4.5.2.1 Total Capital Investment

The plant economic evaluation is based on plant capacity of 500 KTA with operating hours of 8320 hours. The operating rate of 95 % was estimated to give a production capacity of 475 000 tonnes per year.

The total capital investment is estimated to be USD 196 776 836 according to Section 3.3 and the calculated components are tabulated in Table 4.4.

Table 4.4: Estimation of Total Capital Investment

Total Capital Investment	
Production rate factor, F_{PR}	16
Module cost of onsite equipment, C_M, USD	21 844 552
Total bare module cost, C_{TBM}, USD	51 851 604
Direct permanent investment, C_{DPI}, USD	114 073 528
Total permanent investment, C_{TPI}, USD	171 110 292
Total capital investment, C_{TCI}, USD	**196 776 836**

4.5.2.2 Project Cost Estimation

The variable cost and fixed cost are estimated to be a USD 1083/MT and USD 215/MT respectively as shown in Table 4.5.

Table 4.5: Estimation of Production Cost

Variable Cost	
Raw material cost	482 849 484
Catalyst and Chemicals, USD/yr	4 370 000
Utilities Cost	
Electricity, steam, cooling water, USD/yr	12 096 407
Operation (Labor-related) (O)	
Direct wages and benefits (DW&B) , USD/yr	1 747 200
Direct salaries and benefits, USD/yr	262 080
Operating supplies and services, USD/yr	104 832
Technical assistance to manufacturing, USD/yr	360 000
Control laboratory, USD/yr	390 000
Maintenance (M)	
Wages and benefits (MW&B) for fluid handling, USD/yr	5 389 974
Salaries and benefits, USD/yr	1 347 494
Materials and services, USD/yr	5 389 974
Maintenance overhead, USD/yr	269 499
Total Variable Cost, USD/yr	514 576 944
Total Variable Cost, USD/MT	**1083**
Fixed Cost	
Operating Overhead, USD/yr	1 749 350
Property and insurance, USD/yr	2 309 989
Depreciation, USD/yr	
Direct plant, USD/yr	10 362 024
Allocated plant, USD/yr	1 468 437
Rental fees, USD/yr	6 650 000
Licensing fees, USD/yr	13 300 000
General Expenses	66 500 000
Total Fixed Cost, USD/yr	102 339 800
Total Fixed Cost, USD/MT	**215**
Total Production Cost, USD/MT	**1299**

4.5.2.3 Breakeven Point, Return of Investment, Payback Period and Venture Profit Estimation

A chart of breakeven was plotted in Figure 4.11 and the breakeven point was obtained at the point where the sales revenue is equal to the total production cost.

Based on the Figure 4.11, the breakeven point would be at sales revenue of USD 452 430 426/yr and output capacity of 68 %. The rate of return for the plant was estimated to be 20 % with the payback period of 5 years and the venture profit of USD 72 902. The return rate is low as the raw material cost is high which account around 80 % of the production cost. As a result, further study for ethanol production is required in order to reduce the production cost.

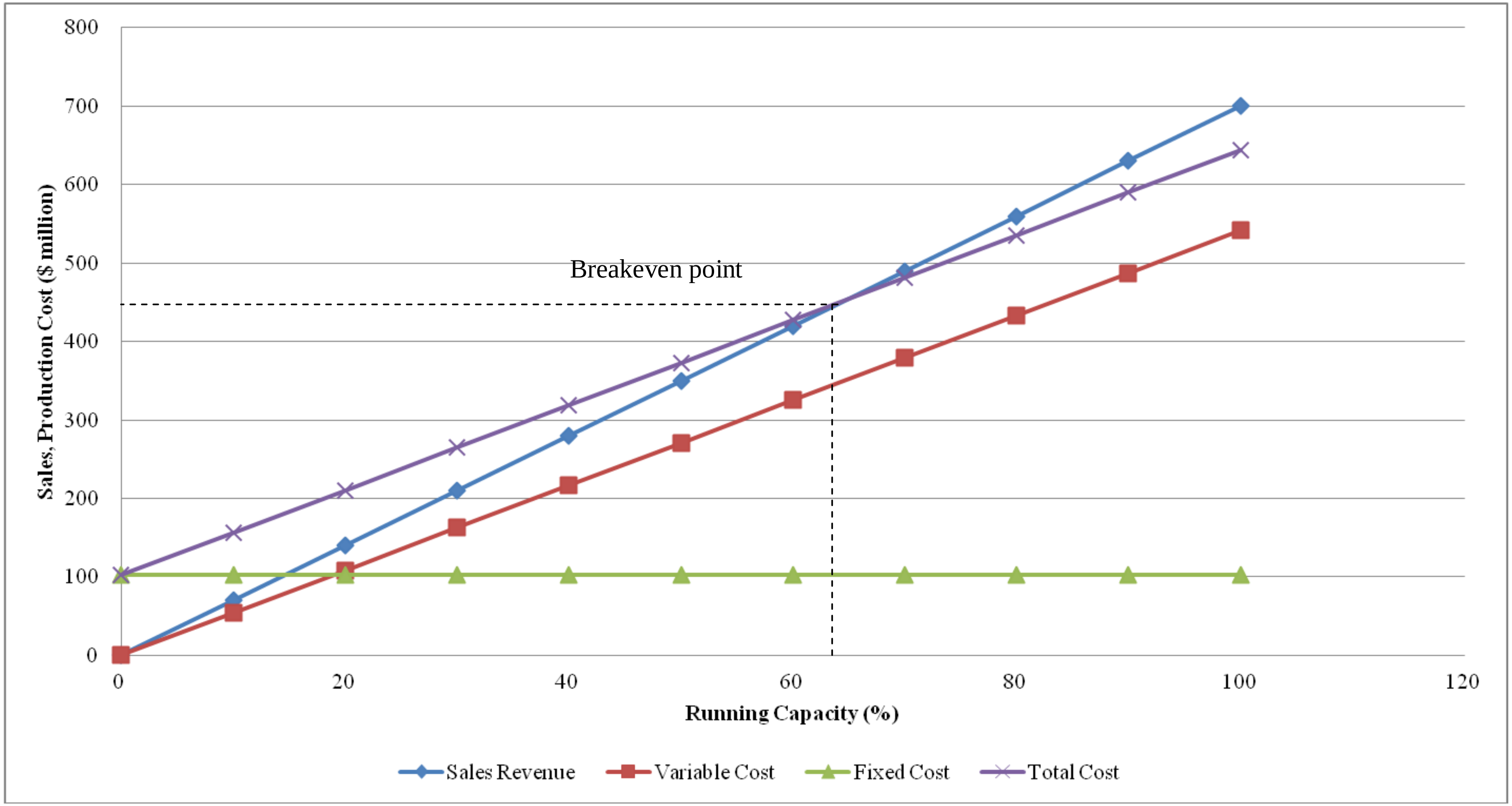

Figure 4.11: Breakeven Point Chart

4.5.2.4 Project Cost Summary

Table 4.6 summarizes the project cost of ETE plant which can help to analyse the economic viability of the project..

Table 4.6: Economic Analysis Review

Components	Results
Plant capacity	500 KTA
Operating rate	95 %
Ethanol price	USD 578/MT
Ethylene selling price	USD 1400/MT
Total capital investment	USD 196 776 836
Variable cost	USD 1083/MT
Fixed cost	USD 215/MT
Contribution ratio	22.62 %
Net profit	USD 39 428 270/yr
Sales profit	USD 48 083 256 yr
Return of investment, ROI	20 %
Payback period, PBP	5 years
Breakeven point	
● **Sales value**	USD 452 430 426/yr
● **Production capacity**	68 %
Venture profit	USD 72 902

4.5.3 Simulation

The dehydration of ethanol to ethylene in a fluidized bed reactor was simulated using HYSYS. In the simulation, UNIQUAC thermodynamic model which estimates the interactions between reaction components was used and the fluidized bed reactor was modelled as conversion reactor.

The simulation of ETE process was based on plant capacity of 500 KTA of ethylene. The HYSYS flow sheet which includes the section of conversion of ethanol to ethylene and purification of ethylene to chemical grade ethylene has shown in Figure 4.12. In the dehydration, liquid ethanol feed (93 wt % ethanol, 7 wt % water) was vaporized in the heater E-100 at 398 ^{o}C. The heated ethanol stream was then introduced into the conversion reactor of CVR-100 at which the conversion of ethanol was set to 99.6 %. The reactor's temperature was maintained at a temperature of 399 ^{o}C using ADJ-2 by adjusting the heat flow. The ethylene and traces of water vapour exit at the top of the reactor.

In the purification section, the top products were cooled to 35 ^{o}C in a cooler to condense some of the water. Removal of water from the cooled ethylene stream was done in an absorber, T-100 where the Water Wash stream was fed at the top stage of the absorber. The bottom product of T-100 is the water stream which contains traceable amounts of ethylene and ethanol. It was introduced into X-100 at which part of the water recovered was recycled to the absorber as Water Wash stream. While the top product of T-100, crude ethylene which contains 99.7 % mole ethylene was passed through a drier unit X-101 to remove residual water. At last, around 100 % mole of ethylene was obtained as a product stream.

The overall mass balances and energy requirements for major components, inlet and outlet streams are shown in Table 4.7, Table 4.8 and Table 4.9 respectively. It can be noted that 100 400 kg/hr of ethanol was required to produce 57 850 kg/hr of ethylene. The energy required by the process was 2.69×10^{8} kJ/hr and the energy removed from the process was 1.74×10^8 kJ/hr. The results were summarized in Table 4.10 and the detailed results are attached in Appendix B.

Table 4.7: Material Balance by Stream

Inlet Material Stream	Mass Flow (kg/hr)	Outlet Material Stream	Mass Flow (kg/hr)
Feed	1.004×10^5	Product	5.785×10^4
Water In	1000	By-product	2093
		Water product	4.146×10^4
Total Inlet Flow	$\mathbf{1.014\times10^5}$	**Total Outlet Flow**	$\mathbf{1.014\times10^5}$

Table 4.8: Material Balance by Component

Component	Inlet Mass Flow (kg/hr)	Outlet Mass Flow (kg/hr)
Ethanol	8.85×10^4	167.86
Water	1.29×10^4	4.34×10^4
Ethylene	-	5.79×10^4
Total	$\mathbf{1.01\times10^5}$	$\mathbf{1.01\times10^5}$

Table 4.9: Energy Balance by Stream

Inlet stream	Energy Flow (kJ/hr)	Outlet stream	Energy Flow (kJ/hr)
Feed	-6.55×10^8	Product	1.08×10^8
Water in	-1.58×10^7	By-product	-2.94×10^7
		Water product	-6.56×10^8
Stream Enthalpy In	$\mathbf{-6.71\times10^8}$	**Stream Enthalpy Out**	$\mathbf{-5.77\times10^8}$
Q-100	1.10×10^4	Q-103	1.74×10^8
Q-101	1.77×10^8	Q-104	-8.36×10^5
Q-102	9.76×10^7		
Q-105	-5.12×10^6		
External Energy In	$\mathbf{2.69\times10^8}$	**Eternal Energy Out**	$\mathbf{1.74\times10^8}$
Total In	$\mathbf{-4.03\times10^8}$	**Total Out**	$\mathbf{-4.03\times10^8}$

Table 4.10: Summary of Simulation Results

Produced ethylene	5.785×10^4
Ethylene/ethanol Yield	0.58 kg/kg
Energy required	2.69×10^8
Energy removed	1.74×10^8

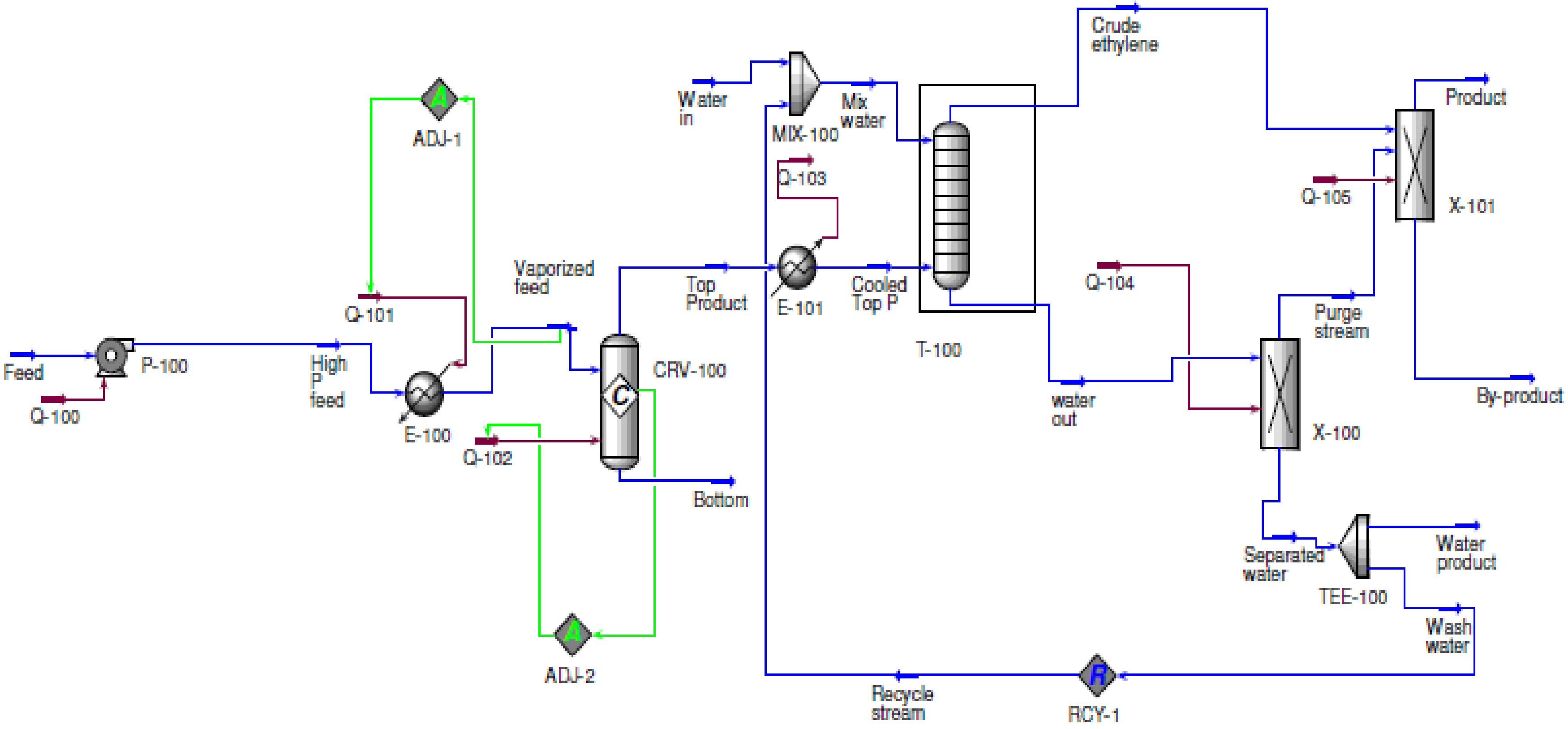

Figure 4.12: Overall Process Design Diagram for Ethylene Production from Dehydration of Ethanol

4.6 Comparison between ETE and Naphtha Cracking

Comparison was made between the steam cracking of naphtha, catalytic cracking of naphtha and ETE processes for ethylene production. They were compared in term of energy use, CO_2 emission, production cost and ethylene yield for petrochemical production.

Based on Table 4.11, steam cracking is the most energy consuming process compared to FCC and ETE. It is the most energy consuming in the chemical industry and emitted nearly 200 million tonnes of CO_2 due to the combustion of fossil fuels in year 2000 (Tao at el., 2008). Cracking provides the heat of reaction and the sensible or latent heat to bring the reactants to the desired reaction temperature which is very high, 750 ℃ to 900 ℃ (Gielen et al, n.d.). It uses maximum energy of 40 GJ per ton of ethylene and emitted around 2 tonnes of carbon dioxide for every ton of ethylene produced which is the highest among the others (Tao et al., 2006).

Besides, the energy use and CO_2 emission for cracking process are higher due to the formation of coke in the cracker. Additional energy is required for decoking or defouling for various sections of a steam cracker. This additional energy use in decoking and defouling is not taken into consideration in the result. With the addition of energy used for shutdown and related maintenance, approximately 1 % to 2 % of total energy used are added (Tao et al., 2006). Therefore, further studies on reducing the coke formation in the steam cracker is required to extend the furnace life and increase the intervals between shutdowns which in turn, reduce the energy consumption and operating cost.

On the other hand, Tao & Patel (2009) have reviewed on Maize starch ETE and Sugarcane ETE process which have integrated with the ethanol production. This leads to significant difference in term of the issues concerned as compared to stand-alone ETE process. However, based on Figure 4.13 and 4.14, it is evident that both of the ETE processes are enjoying a lower energy use and CO_2 emission in ethylene production as compared to naphtha steam cracking process.

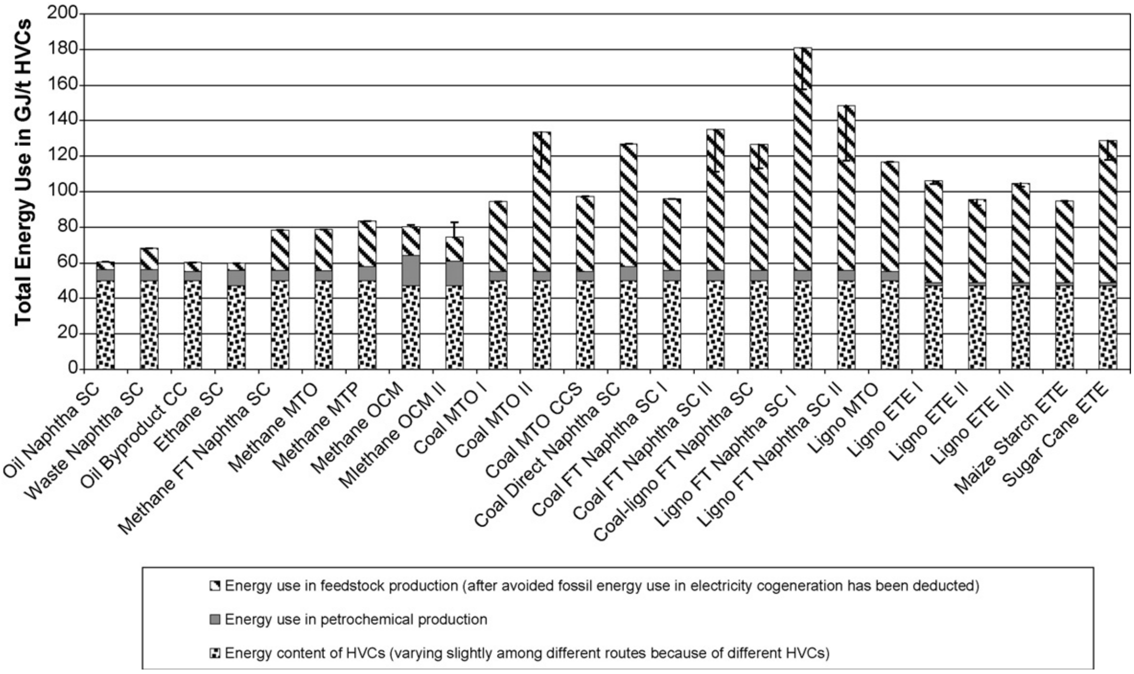

Figure 4.13: Total Energy Used for Different Routes of Ethylene Production

[Source: Tao & Patel, 2009]

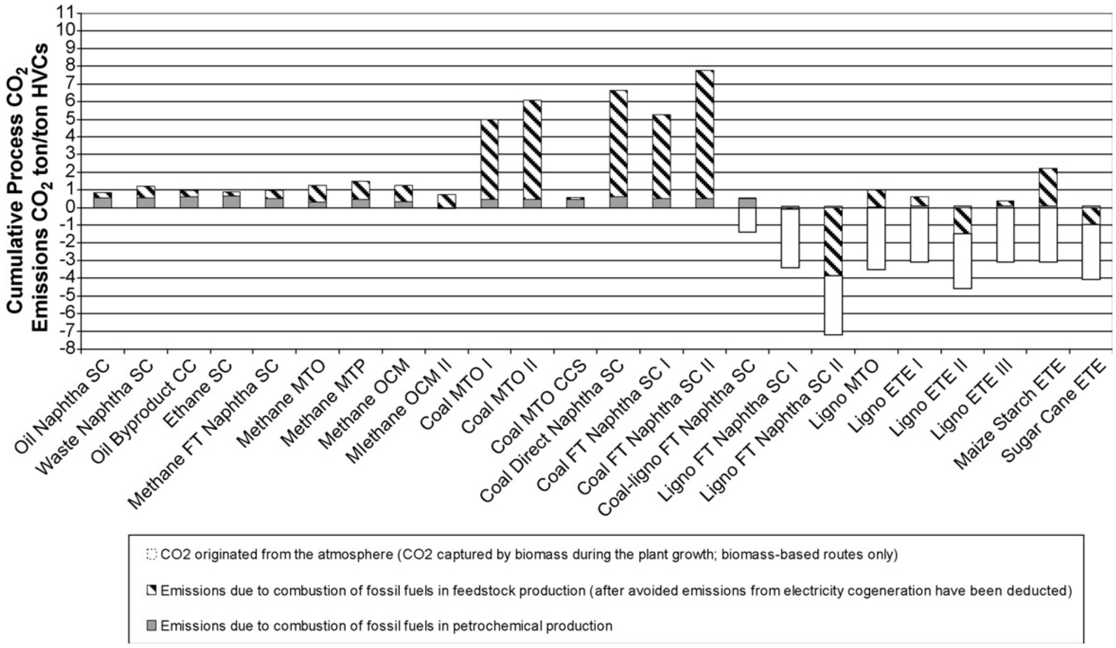

Figure 4.14: CO_2 Emission for Different Routes of Ethylene Production

[Source: Tao & Patel, 2009]

In term of product yield, ethylene yield obtained from ETE routes is higher compared to cracking process as cracking process will produce by-products such as propylene, butylenes and aromatics. The typical ethylene yield obtained from ETE routes is more than 99 wt % which is significantly more than cracking routes.

Besides, the energy cost of naphtha cracking process is high where the energy cost of pyrolysis sector had accounted for 70 % of production cost in typical naphtha-based olefins plants (Tao et al. 2006). In term of the raw material cost, the material cost of naphtha cracking is also higher than ETE process as the naphtha price is USD 1067/MT (Petrochemical Asia, 2012) while the ethanol selling price is USD 690/MT (Agricultural Marketing Resource Center, 2012).

Furthermore, ETE process requires no equipment for distillation leading to lower investment (Kjellin & Johansson, 2010). The technologies such as fluidized bed and adiabatic fixed bed require no external heating fluid enhances the heat transfer and efficiency which in turn, reduces the energy cost and enjoying lower energy cost as compared to naphtha cracking process.

However, the price and production of ethanol suffer considerable seasonal variations, dependent on the geographical location. Brazil which is the world leading ethanol producer whose produces ethanol at very low cost, hence enjoying a significantly lower ethylene production cost, at approximately USD 1200 per ton of bio-ethylene. While higher costs were reported, production based on sweet sorghum at about USD 1700/ton of bio-ethylene in China; production based on corn and sugar beet at about USD 2000/ton and USD 2600/ton in US and European respectively (International Renewable Energy Agency, 2012).

For comparison, the cost of petrochemical ethylene is substantially lower with the global average cost USD 1100/ton of ethylene. The current production cost of bio-ethylene is 1.1 to 2.3 times higher than the global average petrochemical ethylene, thus lignocellulosic bio-ethylene is expected to reduce the gap in the near future (International Renewable Energy Agency, 2012).

Table 4.11: Comparison between Naphtha Cracking and ETE

	Naphtha Steam Cracking	**Catalytic Naphtha Cracking (FCC)**	**Ethanol-to-Ethylene**	
			Maize Starch ETE	**Sugarcane ETE**
Energy use (GJ/ton ethylene)	26–31 (typical)	19	-	-
Energy use (GJ/ton HVCs)	14–17	10	37 (Energy use in the feedstock and petrochemical production)	-17 (Energy use in the feedstock and petrochemical production)
Ethylene Yield (wt %)	29–34	38	> 99	> 99
CO_2 emission (ton CO_2/ton ethylene)	1.8–2.0	-	-	-
CO_2 emission (ton CO_2/ton HVCs)	1.6–1.8	-	Lower	Lower
Reference	Tao et al., 2006		Tao & Patel, 2009	

4.7 Commercial ETE Plants and Research Activities

Given the public concern about global warming, scarcity of fossil fuels and increasing oil prices, bio-based polyethylene has become increasingly attractive today. Besides, customer interest in renewably sourced materials, particularly in the packaging area, is the primary demand driver for bio-based polyethylene. Customers have started to exert pressure on polymer-producing companies especially polyethylene producers to deliver sustainable solutions where renewable feedstock is an important factor. It is inevitable that there is a large potential market demand for bio-ethylene. The following companies have planned to produce bio-ethylene from bio-ethanol at certain capacity:

- Braskem Ethanol-to-Ethylene Plant, Brazil

Braskem, a Brazilian petrochemicals company has inaugurated the first large scale ethylene plant in the Rio Grande do Sul state of Brazil in September 2010. The production capacity of 200 000 tonnes per year uses 100 % renewable raw materials which is the sugarcane based ethanol as feedstock (Net Resources International, 2011). In addition, Braskem planned to establish the second factory to produce polyethylene from sugarcane-based ethanol, once the first plant commenced operation in October, 2010.

- Dow Chemical Company and Crytsalsev, Brazil

Dow Chemical Company, the world’s largest producer of polyethylene has joint venture with Crystalsev, one of Brazil largest ethanol producers to produce bio-based polyethylene at annual production capacity of 350 000 tonnes in 2012 (EnvaPack, 2007).

- Solvay, Brussels

The Solvay group, one of the world’s leading vinyls producer had created an integrated plant to produce ethylene with ethanol originating from sugarcane. The purpose for this investment is to expand and increase the competiveness of its vinyl production plant in Santo Andre, Brazil. This expansion of Santo Andre plant was estimated to be completed by 2010 with annual installed capacity of 360 000 tonnes of polyvinylchloride (PVC) and 60 000 tonnes sugarcane based bio-ethylene (BioPact, 2007).

- Borealis, Austria

Borealis is an organisation which focuses on producing plastic materials using the feedstocks of naphtha, butane, propane and ethane. Due to the existing production capacities are unable to meet the demand, Borealis had planned to expand its production of ethylene at Stenungsund, Sweden, by using bio-ethanol as a raw material (Nexant, 2009).

- Total, IFPEN and Axens

Total Petrochemicals, IFP Energies nouvelles (IPFEN) and its subsidiary Axens have entered into a partnership to develop new technology that can produce bio-ethylene by dehydrating ethanol. The technology is expected to be available for industrial implementation by the end of 2011 (Natural Resources International, 2011).

- Greencol Taiwan Corporation Bio-MEG Plant, Taiwan, China

Greencol Taiwan Corp., a joint venture between China Man-Made Fiber Corp. and Toyota Tsusho Corp. has planned to build a 100 000 tonnes bio-based ethylene plant in Kaoshiung, Taiwan (Natural Resources International, 2010). The feedstock used will be the sugarcane based ethanol supplied by Petrobras.

- China’s Companies

Sinopec has planned to expand a 7000 tonnes per year bio-ethylene experiment unit at Sichuan Vinylon Works. Another company in China, Anhui BBCA Biochemical Co has included one million tonnes per year of bio-ethylene project in its 11th Five-Year Plan (2006-2010). Besides, according to Jilin government official, Songyuan Ji’an Biochemical Group Co is also considering a project to produce 300 000 ton/year of ethylene from corn (ICIS, 2005).

CHAPTER 5

CONCLUSION AND RECOMMENDATIONS

5.1 Conclusion

Lummus fluidized bed technology combined with the HZSM-5 zeolite catalyst was found to be the better technology compared to the other technologies as it gives almost 100 % of ethanol conversion, ethylene selectivity and ethylene yield at the average temperature. This is due to the use of fluidized bed reactor which enables the operating temperature to be controlled and optimum temperature can be obtained. Besides, fluidized bed reactor allows continuous regeneration of HZSM-5 catalyst which permits the catalyst activity to be held constant.

Besides, through the simulation, the results showed that 100 400 kg/hr of ethanol was required to produce 57 850 kg/hr of ethylene which is equivalent to 0.58 kg ethylene/kg ethanol. The energy required by the process was 2.69×10^8 kJ/hr and the energy removed from the process was 1.74×10^8 kJ/hr.

Apart from that, the investment cost for ETE plant was estimated to be around USD 197 million which is lower than that of naphtha cracking plant as equipment of distillation column is not required for production. However, the production cost of 1299/ton of ethylene produced through ETE route was estimated which is much higher than that of equivalent petrochemical ethylene. This is due to the high ethanol cost which accounted about 80 % of total production cost. In such case, the ROI of 20 % and payback period of 5 years were obtained.

As a conclusion, ETE process is viable as some of the companies are going to commence the production of ethylene through ETE especially at the locations such as Brazil which have low cost ethanol. It is more environmental friendly than the naphtha cracking as it emits less greenhouse gases and consumes less energy for ethylene production. Besides, ethanol as the feedstock for ETE process can be easily produced by simple conversion of agricultural products and the production capacity is increasing due to the mature technologies enhances the feasibility of ETE process. However, the production cost of ethylene through ETE process is higher than that of the equivalent petrochemical ethylene. Thus, development of cheap and sustainable conversion processes of low cost lignocellulosic biomass is essential in the future.

5.2 Recommendations

Stand-alone dehydration plant that purchases ethanol and sells ethylene at normal market prices would be only marginally profitable. Economics would improve if the ethylene plant could be integrated with the upstream ethanol plant. Thus further study on the effects of integrating the ethylene production with the upstream ethanol production from lignocellulosic feedstock is recommended in order to overcome the drawbacks of stand-alone ETE process which having high production cost.

Based on the situation currently, ETE process is not highly recommended as the ethanol prices are varied seasonally. Thus, ethylene production via co-processing of ethanol and hydrocarbons through FCC is suggested for further study which allows the supply of ethanol to be interrupted when processing thereof is not economically attractive.

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APPENDICES

APPENDIX A: Ethanol Production in South East Asia

Table A.1: Ethanol Production in Thailand

[Source: Liumsirijarern, 2011]

Ethanol Plant in Year 2011		
Ethanol Plant	**Main Raw Material**	**Capacity (litre/day)**
Porn Vilai International Trading Group	Molasses	2,500
Thai Agro Energy Public	Molasses	150,000
Thai Alcohol	Molasses	200,000
Kon Kaen Alcohol	Molasses	150,000
Thai Nguan Ethanol	Cassava Root	130,000
Thai Sugar Ethanol	Molasses	100,000
K.I. Ethanol	Molasses	100,000
Petro Green (Kalasin)	Molasses	100,000
Petro Green (Pukheaw)	Molasses	200,000
Eakkarat Pattana	Molasses	200,000
Thai Rung Ruang Energy	Molasses	120,000
Ratchaburi Ethanol	Cassava Chip	150,000
E.S.Power	Molasses	150,000
Maesod Palungngan SA.AD	Sugarcane Juice	200,000
Subtip Ethanol	Cassava Chip	200,000
Tai Ping Ethanol	Cassava Root	150,000
Psc Starch Production	Cassava Chip	150,000
Petro Green (Danchang)	Molasses	200,000
Khon Kaen Alcohol (Bo Phloi)	Molasses	150,000
Ethanol Plant Under Construction		
TPK Ethanol (Phase 1)	Cassava Chip	340,000
TPK Ethanol (Phase 2,3)	Cassava Chip	680,000
Double A Ethanol	Cassava Chip	250,000
Sima Interproduct	Cassava Chip	150,000
Impress Technology	Cassava Chip	200,000
Thai Agro Energy Public	Cassava Chip	200,000

Table A.2: Ethanol Production in Indonesia

Ethanol Plant	Feedstock	Capacity (million litre/year)	Source
Existing Facilities in 2008			
Anugrah Kunia Abadi & BPPT	Cassava	2.5	Zarrah Duniani & Mas Putra Agung, 2009
Medco Ethanol	Cassava	270	
Molindo Raya	Molasses	150	
Sugar Group	Molasses	530	
Expected Facilities in 2010			
Sampoerna	Cassava	600	Panaka & Yudiarto, 2007
Indo Lampung	Molasses	150	
Salim Group	Sugar	70	Dillon, Laan, & Dillon, S., 2008
Sungai Budi	Cassava	120	
Berlian Energy	Sugar	1.8	
Wilmar Group	Sugar	70	
Satria, Broenzeoak	Sugar	300	
Mitsui Petrobras	Sugar	500	
SORINI TBK	Cassava	200	
Angel Product	Sugar	10	
EN3 Korea	Cassava	180	
Small Producers for Pertamina	Sugar	200	
Community Project Scale	n.a.	600	

Table A.3: Ethanol Production in Philippine

[Source: Corpuz, 2009]

Ethanol Plant in 2008		
Investor	**Feedstock**	**Capacity (millions litre/year)**
Bronseoak, San Carlos Bioenergy	Sugarcane and Molasses	30
Leyte Agri Corp.	Molasses	9
Ethanol Plant in 2009		
Roxol Bio-energy	Molasses	30
Cavite Bio-Fuel	Sugarcane and Molasses	37.5
Ethanol Plant in 2010		
Fuel Inc.	Sugarcane and Molasses	30
Bio-Fuel Int	Sugarcane/Sweet Sorghum/Molasses	38
Negros Biochem	Sugarcane and Molasses	30
Bronzeok, Southern Bukidnon	Sugarcane and Molasses	45
Bronzeoak, Capas	Sugarcane and Molasses	30
JGC/ITOCHU	Sugarcane and Molasses	30
JG Summit	Molasses	30
Bio fuels 88	Molasses	18
North Cotabato	Molasses	n.a.
Robson Agro Ventures	Cassava	60
Alson’s Power	Cassava	30
Eastern Petroleum	Cassava	30
Ethanol Plant in 2011		
Basic Energy	Sugarcane/ Molasses/ Cassava	30
E-cane Fuel	Sugarcane and Molasses	45
Daebong LS	Molasses	30
Interested Party	Sugarcane/ Molasses	30
Palawan Bio Energy	Sugarcane and Molasses	30

Table A.4: Ethanol Production in Vietnam

[Source: New Energy Foundation, n.d.]

Ethanol Plant under Operation	
Company	**Capacity (million litre/year)**
Dong Xanh Joint Stock Company	125
Tung Lam Company Ltd.	70
Ethanol Plant under Construction	
Dai Viet Joint Stock Company	50
Tan Phat Joint Stock Company	50
Ethanol Plant Expected to start Operation in 2011	
Joint Stock Company .Petrol.& Biofuel	100
Midlands Central biofuel Joint Stock Company	100
VN Oil Company	100
Dakto Bio-ethanol factory	65
Ethanol Plant Expected to start Operation in 2012	
Thao Nguyen Joint Stock Company	100
Ethanol Plant Expected to start Operation in 2013	
Thai Viet Joint Stock Company	62.7

Table A.5: Ethanol Production in Myanmar

[Source: New Energy Foundation, n.d.]

Ethanol Plant under Operation	
Ethanol Plant	**Capacity (gal/day)**
Ethanol Distillery No.2 Sugar mill (MSE)	500
Kantbalu Distillery (MEC)	3,000
Mattaya Distillary	15,000
MaungKone	37,500
Pyinhtaunglay	45,000
Ethanol Plant under Construction	
Taungsinaye Distillery (MEC)	3,000

APPENDIX B: Simulation Results