BATCH SCHEDULING AND ECONOMIC ANALYSIS OF 200,000 LITERS PER DAY BIODIESEL PRODUCTION

MISS KANYAPORN LERTWIMOLKASEM

A SPECIAL RESEARCH PROJECT SUBMITTED IN PARTIAL FULFILLMENT
OF THE REQUIREMENTS FOR
THE DEGREE OF MASTER OF ENGINEERING (CHEMICAL ENGINEERING)
FACULTY OF ENGINEERING
KING MONGKUT'S UNIVERSITY OF TECHNOLOGY THONBURI
2011

Batch Scheduling and Economic Analysis of 200,000 Liters Per Day Biodiesel Production

Miss Kanyaporn Lertwimokasem B.Eng. (Chemical Engineering)

A Special Research Project Submitted in Partial Fulfillment

of the Requirements for

The Degree of Master of Engineering (Chemical Engineering)

Faculty of Engineering

King Mongkut's University of Technology Thonburi

2011

Special Research Project Committee

......V. Loha...... (Asst. Prof. Veara Loha, Ph.D.)	Chairman of Special Research Project Committee
.... (Assoc. Prof. Thongchai Srinophakun, Ph.D.)	Member and Special Research Project Advisor
......N. Nimitsiriwat...... (Nonsee Nimitsiriwat, Ph.D.)	Member

Special Research Project Title	Batch Scheduling and Economic Analysis of 200,000 Liters Per Day Biodiesel Production
Special Research Project Credits	6
Candidate	Miss Kanyaporn Lertwimolkasem
Special Research Project Advisor	Assoc. Prof. Dr. Thongchai Srinophakun
Program	Master of Engineering
Field of Study	Chemical Engineering
Department	Chemical Engineering
Faculty	Engineering
B.E.	2554

Abstract

This thesis is to perform and to simulate the production of 200,000 liters per day of biodiesel by a batch process. The study covers both steady-state and batch simulations. Feed stock of palm oil with weight fractions of triolein, tripalmitin and trilinolein are 0.4019, 0.4738 and 0.1123; respectively. The transesterification using KOH as a catalyst is the main reaction. Batch time and liquid volume calculated from batch simulation were used in the economic section in order to calculate the payback period. In this study, addition glycerol purification section is introduced to study by varying the number of mixers from 1 to 5, reactors from 1 to 3 and washing tanks from 1 to 2. The result showed that the process with glycerol purification section has a lower payback period than without this section even though the additional glycerol purification section needs a higher investment cost. Their two products can make the payback period shorter. Then the crude palm oil price is investigated. The study revealed that the more expensive the palm oil price, the longer payback period is; consequently, the process with purification section can absorb higher palm oil price than process without the purification section. The amount of impurity is also interested. This research considered only on a beta carotene which affects the colour of glycerol and palmitic acid. This section studied two reactions transesterification as a main reaction and saponification as a side reaction. The composition in feed stock includes triolein, tripalmitin, trilinolein, beta carotene and fatty acid. The composition of fatty acid was varied from 0, 2.5, 7.5 and 12.5 % weight with content beta carotene at 2.5 % weight. The result shows that 12.5% impurity gives the highest payback period, followed by 8.5, 2.5 and 0% impurity at the lowest payback period. In case of the impurity in feedstock is between 2.5 to 7.5%, it appears to only slightly affect the payback period. In conclusion, the biodiesel process with glycerol purification section consists of 1 mixer, 1 reactor and 1 washing tank is optimum for investment in biodiesel production.

Keywords: Batch process/ Biodiesel/ Triolein/ Tripalmitin / Trilinolein/ Palmitic acid

หัวข้อโครงการศึกษาวิจัย	การจัดการผลิตแบบกะและวิเคราะห์ทางเศรษฐศาสตร์ของกระบวนการผลิตไบโอดีเซลขนาด 200,000 ลิตรต่อวัน
หน่วยกิต	6
ผู้เขียน	นางสาว กันยาพร เลิศวิมลเกษม
อาจารย์ที่ปรึกษา	รศ.ดร. ธงไชย ศรีนพคุณ
หลักสูตร	วิศวกรรมศาสตรมหาบัณฑิต
สาขาวิชา	วิศวกรรมเคมี
ภาควิชา	วิศวกรรมเคมี
คณะ	วิศวกรรมศาสตร์
พ.ศ.	2554

บทคัดย่อ

วัตถุประสงค์ของงานวิจัยนี้คือการออกแบบกระบวนการผลิตไบโอดีเซลแบบกะกำลังการผลิต 200,000 ลิตรต่อวัน ในงานวิจัยนี้จะทำศึกษาทั้งระบบที่สภาวะคงตัวและระบบแบบกะ โดยน้ำมันปาล์มที่ใช้เป็นสายป้อนจะประกอบด้วยไตรโอเลอินร้อยละ 40.19 ไตรปาล์มมิทินร้อยละ 47.38 และไตรไลโนเลอินร้อยละ 11.23 โดยน้ำหนัก ปฏิกิริยาทรานเอสเทอริฟิเคชันโดยใช้โพแทสเซียมไฮดรอกไซด์เป็นตัวเร่งปฏิกิริยาเป็นปฏิกิริยาเคมีหลักในการผลิตไบโอดีเซล ในการศึกษานี้จะเปรียบเทียบกระบวนผลิตไบโอดีเซลแบบเพิ่มและไม่เพิ่มกระบวนการเพิ่มความบริสุทธิ์ของกลีเซอลีน ซึ่งจะใช้กระบวนการผลิตหลักที่ประกอบด้วยถังกวน ถังปฏิกรณ์เคมีและถังล้าง เหมือนกัน นอกจากนี้ยังมีการศึกษาผลของจำนวนเครื่องถังกวน ถังปฏิกรณ์เคมีและถังล้าง โดยทำการเพิ่มและลดจำนวนเครื่องดังนี้ 1 ถึง 5 ถังกวน 1 ถึง 3 ถังปฏิกรณ์เคมีและ 1 ถึง 2 ถังล้าง จากผลการทดลองพบว่ากระบวนผลิตที่เพิ่มส่วนของการเพิ่มความบริสุทธิ์ของกลีเซอลีนจะมีระยะเวลาการคืนทุนเร็วกว่ากระบวนการผลิตที่ไม่เพิ่มส่วนดังกล่าว เป็นเพราะ แต่การเพิ่มส่วนเพิ่มความบริสุทธิ์ของกลีเซอลีนก็ต้องใช้เงินลงทุนที่สูงขึ้นโรงงานนี้สามารถขายกลีเซอลีนได้เพิ่มขึ้นอีกหนึ่งผลิตภัณฑ์ นอกจากนี้ยังมีศึกษาผลกระทบของราคาน้ำมันปาล์มซึ่งพบว่าถ้าราคาน้ำมันปาล์มสูงขึ้นจะทำให้ระยะเวลาการคืนทุนนานขึ้น กระบวนการผลิตที่เพิ่มกระบวนการเพิ่มความบริสุทธิ์ของกลีเซอลีนสามารถลองรับผลกระทบจากการเพิ่มของราคาน้ำมันปาล์มได้มากกว่ากระบวนการที่ไม่มีการเพิ่มส่วนดังกล่าว ผลการทดลองข้างต้นเป็นใช้เพิ่งไตรโอเลอิน ไตรปาล์มมิทินและไตรไลโนเลอินเป็นองค์ประกอบในน้ำมันปาล์มซึ่งในความเป็นจริงมักมีสารเจือปนด้วย ในงานวิจัยนี้จึงทำการศึกษาผลของสิ่งเจือปนและ

จะสนใจเพิ่งแบต้าแคโรทีนซึ่งมีผลต่อสีของกลีเซอลีนที่ขาย และ กรดไขมันอิสระกรดปาล์มมิติกเท่านั้น ในส่วนของการศึกษาผลของสิ่งเจือปนนั้นจะมีสองปฏิกิริยาเคมีเกิดขึ้นในการผลิตไบโอดีเซลคือปฏิกิริยาทรานเอสเทอริฟิเคชันและปฏิกิริยาสปอนนิฟิเคชันซึ่งเป็นปฏิกิริยาข้างเคียง องค์ประกอบของสายป้อนจะประกอบด้วยไตรโอเลอิน ไตรปาล์มมิทิน ไตรไลโนเลอิน แบต้าแคโรทีนและกรดไขมันอิสระ โดยที่จะเปรียบเทียบร้อยละของกรดไขมัน 0, 2.5, 7.5 และ 12.5 โดยน้ำหนัก ขณะที่ให้ร้อยละของแบต้าแคโรทีนคงที่ที่ 2.5โดยน้ำหนัก จากผลการทดลองพบว่าสารป้อนที่มีกรดไขมันอิสระ 12.5% ต้องใช้ระยะเวลาการคืนทุนนานที่สุด ลองลงมาคือ 7.5% และ 2.5% ซึ่งให้ค่าใกล้เคียงกัน และ 0% ใช้ระยะเวลาการคืนทุนน้อยที่สุด นั่นหมายความว่าถ้าในสารป้อนมีสิ่งเจือปนในช่วง 2.5% ถึง 7.5% จะไม่ค่อยมีผลกระทบต่อระยะเวลาการคืนทุน แต่ถ้าสารป้อนมีสิ่งเจือปนมากกว่า 7.5% ปริมาณสิ่งเจือปนจะมีผลต่อระยะเวลาการคืนทุนมาก นั่นคือถ้าปริมาณสารเจือปนเพิ่มเพียงเล็กน้อย อาจส่งผลให้ระยะเวลาการคืนทุนนานขึ้นอีก ดังนั้นจึงสรุปว่ากระบวนการผลิตไบโอดีเซลที่เพิ่มกระบวนการเพิ่มความบริสุทธิ์ของกลีเซอลีน มีจำนวนถังกวน 1 ถัง จำนวนถังปฏิกรณ์เคมี 1 ถัง และจำนวนถังล้าง 1 ถัง เหมาะสมแก่การลงทุนมากที่สุด

คำสำคัญ: กระบวนการผลิตแบบกะ/ ไบโอดีเซล/ ไตรโอเลอิน/ ไตรปาล์มมิทิน/ไตรไลโนเลอิน/
กรดปาล์มมิติก

ACKNOWLEDGEMENTS

I would like to express my thanks to Assoc. Prof. Dr. Thongchai Srinophakun, my advisor for support, guidance and encouragement. This project cannot be completed without Dr. Adam Harvey and Dr. Jonathan Lee at Newcastle University for their valuable suggestions and comments during developments of this oversea study. My appreciation also goes to Asst. Prof. Dr. Veara Loha and Dr. Nonsee Nimitsiriwat, menbers of committees, for their suggestions and comments.

Moreover, I would like to thanks KMUTT, KU laboratory and ChEPS where give me a chance to study a lot of knowledge and experience. My classmates helped and suggested me when I have problem. Finally, I have to say thank you to my family who supports and understands everything in my life.

CONTENTS

LIST OF TABLES

TABLE **PAGE**

TABLE **PAGE**

LIST OF FIGURES

FIGURE **PAGE**

FIGURE **PAGE**

NOMENCLATURES

A	=	Arrhenius constant
ASTM	=	The American Society for Testing and Materials
^oC	=	Degree Celsius
C	=	Total product cost
$C_{Building}$	=	Cost of building
C_{IBM}	=	Total cost of installed equipment
C_{Land}	=	Cost of land
$C_{Offside}$	=	Cost of offside
C_{Royal}	=	Cost of royalties
$C_{Startup}$	=	Cost of plant start up
COM	=	Cost of Manufacturing
D	=	Depreciation
DG	=	Diglyceride
E	=	Activation energy
EN	=	The European Norm
FCI	=	Fixed Capital Investment
GE	=	General Expense
GL	=	Glycerol
k(T)	=	Kinetic rate of reaction
M	=	Maintenance
MG	=	Monoglyceride
MACRS	=	Modified Accelerated Cost Recovery System
MW&B	=	Maintenance Wages and Benefit
O	=	Labor-related operation
R	=	Gas Constant
T	=	Temperature
TCI	=	Total Capital Investment
TDC	=	Total Direct Cost
TG	=	Triglyceride
WC	=	Working Capital

CHAPTER 1 INTRODUCTION

1.1 Background and Motivation

The diesel demand in the worl is increasing while petroleum fuel oil is finite and non-renewable. Moreover, the rising high cost of diesel and the diesel combustion contribute many problems seed a global worming. Biodiesel can offer as a substitute for petroleum based diesel (Meher *et.al.,* 2004; Fukuda *et.al.,* 2001). Biodiesel is one of the alternative sources which can use in diesel engines and is renewable. In addition to this, it decreases the emission of CO_x, SO_x and unburned hydrocarbons in the combustion process (U.S. Department of Energy, 2006). Comprise of petroleum-based diesel and biodiesel, biodiesel can reduce global warming, it is safer to transport and handle because of its higher flash point (great than 150^{o}C) than traditional petroleum (77^{o}C), and is non-corrosive for human contact (http://www.terrabioenergy.com/benefits). It also has more lubricating than diesel fuel which is better for engines (Zuhair, 2007).

There are many methods to produce biodiesel such as direct use and blending, microemulsions, pyrolysis/thermal cracking and transesterification (Ma and Haa, 1999). This work focuses on biodiesel production by transesterification because it is the more commonly used method (Ma and Haa, 1999). Transesterification is the reaction between vegetable oils such as palm oil, coconut oil, soy bean oil, ground nut oil, castor oil, sesame oil, sunflower oil and Jatropha oil, animal fats or waste vegetable oil with alcohol to produce ester and glycerol by using a catalyst. There are 2 kinds of catalyst; homogeneous and heterogeneous catalyst. Homogeneous catalysts normally use a strong base for example NaOH and KOH while heterogeneous catalysts use a variety of bases e.g. Lipase, ZrO_2-SiO_2 and SrO. Both homogeneous and heterogeneous catalysts have advantages and disadvantages. Homogeneous catalysts have a high conversion and fast reaction rate but has saponification as a side reaction and it is difficult to separate the biodiesel and catalyst.

Scheduling is a critical issue in operations process and is crucial for improving production performance. It is a decision making process to determine the locations, times and sequences for processing activities with finite units and resources to achieve certain objectives, such as minimize the payback period. This research is interested in the design and simulating of batch process for biodiesel production in chemical plants and economic analysis.

1.2 Objectives

1.2.1 To simulate the conventional distillation of biodiesel production rate 20,000 liters/day with the biodiesel specification of EN 14214 and ASTM D 675.

1.2.2 To study the effect of beta-carotene and palmitic acid as the impurity.

1.2.3 To propose the optimum of unit operation.

1.2.4 To perform the sensitivity of crude palm oil price.

1.3 Scope of Work

1.3.1 The simulation cover only transesterification and saponification reaction by conventional distillation with using KOH (homogeneous catalysts) for biodiesel production.

1.3.2 Beta-carotene and fatty acid as impurities in feed stream.

1.3.3 The product of bidiesel from palm oil has mainly their components; tri-palmitic, tri-linoleic and tri-oleic.

1.3.4 The number of unit operations are 1 either 5 mixer tanks, 1 either 3 reactors and 1 or 2 washing tanks in first and second washing step.

1.3.5 Payback period is the only concerning economic analysis that concerned.

1.3.6 ASPEN PLUS and ASPEN Batch simulation are used.

1.3.7 In batch simulation, reactor, mixing and settling tank were batch operating units meanwhile pump, exchanger and distillation were continuous operating units.

1.3.8 Resident time of batch unit is assumed to be the base case one.

1.3.9 Utility selection followed Seider and economic calculation followed work of Towler.

1.4 Expected Results

This study can be applied in biodiesel production plant and other plants which are batch process and learn about Gantt chart which is important in scheduling.

CHAPTER 2 THEORY AND LITERATURE REVIEW

This chapter presents the related literatures and theories. The first part is biodiesel such as history, reaction, biodiesel standard and production process. The second is the understanding of batch process. Economic analysis comes in the third part. Next, it is literature review.

2.1 Biodiesel

2.1.1 History of Biodiesel

Duffy and Patrick first introduce the transesterification of vegetable oil in 1853 (Bell *et al*. 2007). In 1901, Rudof Diesel presented the first biodiesel engine at the World's Exhibition in Paris. In that model engine, he compressed peanut oil in ignition engine. Vegetable oil was used as fuel in diesel engine until around the 1920s thereafter petroleum diesel quickly become the fuel of choice for the diesel, since it was cheaper to produce (Bell *et al*. 2007). In the middle 1970s petroleum fuel crisis, vegetable oil and their alternatives were revived invert in developing biodiesel as an alternative to petroleum diesel (http://www.plantoils.eu/palm_oil.html). Recently, because of the increasing in crude oil prices, limited resources of fossil oil and environmental concerns, the biodiesel become a promising alternative fuel to replace the tradition petroleum diesel.

2.1.2 Potentials of Biodiesel Production in Thailand

Raw material or feedstock with potential for biodiesel production in Thailand are both the used wasted vegetable cooking oil and the virgin vegetable oil. This oil include Palm oil, Coconut oil, Soy bean oil, Ground nut oil, Castor oil, Sesame oil, Sunflower oil and Jatropha oil. The properties of biodiesel depend on number of carbon atom and number of double bound triglyceride as seen in Figure 2.1.

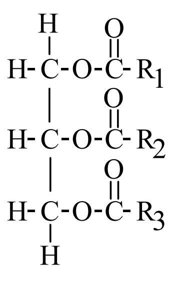


Figure 2.1 R1, R2 and R3 are hydrocarbon chain of fatty acyl groups of triglyceride

2.1.3 Palm oil

Palm oil is quantitatively the highest production yield and increasing plantation area over other major oil crops in Thailand. Chemical compositions and properties of both crude palm oil and crude palm kernel are suitable as biodiesel's feedstock because palm oil and palm kernel oil are composed of fatty acids, esterified with glycerol just like any ordinary fat. Both are high in saturated fatty acids, about 50% and 80%, respectively. The palm oil fatty acids composition are shown in Table 2.1.

Table 2.1 Fatty acid compositions of palm oil

Fatty acid	Percent
Palmitic acid (16:0)	44.3
Myristic (14:0)	1.0
Stearic acid (18:0)	4.6
Oleic acid (18:1)	38.7
Linoleic acid (18:2)	10.5
Other/unknown	0.9
Total saturated fatty acids	50
Total unsaturated acids	50

2.1.4 Biodiesel Standard

The parameters according to standard of pure biodiesel (B100) must be satisfied before it can be referenced as a pure fuel or being blended with petroleum-based diesel fuel. Biodiesel, B100, specification (ASTM D 6751–02 requirements) and the EN 14214 is an international standard that describes the minimum requirements for biodiesel produced from biodiesel fuel stock (also known as fatty acid methyl esters). There are two international standards; EN 14212 and ASTM D 6751 as shown in Table 2.2. (Demirbas, 2008; Halek, 2009)

Table 2.2 The biodiesel specification of EN 14214 and ASTM D 6751

Properties	EN 14214		ASTM D 6751	
	Unit	Limits	Unit	Limits
Ester content	% (m/m)	96.5	-	-
Density at 15 °C	kg/m^3	860-900	-	-
Viscosity at 40 °C	mm^2/s	3.5-5.0	mm^2/s	1.9-6.0
Flash point	°C	120 min	°C	130 min
Sulfur content	mg/kg	10 max	% mass	0.05 max
Carbon residue	% (m/m)	0.3 max	% mass	0.05 max
Cetane number		51 min		47 min
Sulfated ash	% (m/m)	0.02 max	% mass	0.02 max
Water content	mg/kg	500 max	% volume	0.05 max
Total contamination	mg/kg	24 max	-	-
Copper strip corrosion	rating	class 1		No. 3 max
Cloud Point	-	-	°C	Report
Oxidation stability	hours	6 min	-	-
Acid Value	Mg KOH/g	0.5 max	Mg KOH/g	0.8 max
Iodine value	g/100g	120 max	-	-
Linoolenic acid ME	% (m/m)	12 max	-	-
Polyunsat ME	% (m/m0	1 max	-	-
Methanol content	% (m/m)	0.2 max	-	-
Monoglyceride	% (m/m)	0.8 max	-	-
Diglyceride	% (m/m)	0.2 max	-	-
Triglyceride	% (m/m)	0.2 max	-	-
Free glycerol	% (m/m)	0.02 max	% (m/m)	0.02
Total glycerol	% (m/m)	0.25 max	% (m/m)	0.24
Alkali metals (Na+K)	mg/kg	5 max	-	-
Phosphorus content	mg/kg	10 max	% mass	0.001 max
Distillation temp.	-	-	°C	360 max

2.1.5 The production of Biodiesel

There were a number of methods for the product of biodiesel namely direct use and blending, microemulsions, pyrolysis/thermal cracking and transesterification (Ma and Haa, 1999). Direct use and blending is the use of pure vegetable oils or the blending of vegetable oil and diesel fuel or solve or ethanol in various ratio (Khan, 2002; Srivastava and Prasad, 2000). This method has been generally considered to be not satisfactory and impractical for biodiesel and indirect injection engines because of injector coking and trumpet formation, more carbon deposits, oil ring sticking and thickening, and gelling of the engine lubricant oil (Demirbas, 2002). Microemulsions are defined as a colloidal equilibrium dispersion of optically isotropic fluid microstructure, with dimension generally in the range of 1 to 150 nm. These are formed spontaneously from two normally immiscible liquids and one or more ionic or non-ionic amphophiles. A microemulsion with simple alcohol such as methanol and 1-butanol has been studied as a means of solving the problem of high viscosity of vegetable oils. However, significant injector needle sticking, heavy carbon deposits, incomplete combustion and an increase of lubricating oil viscosity were reported in long term testing (Fukuda *et al*., 2001). Pyrolysis/thermal cracking refers to the chemical changes caused by the application of thermal energy in the presence of air or nitrogen sparge (Fukuda *et al*., 2001). In some situation, this is with the aid of a catalyst leading to the cleavage to of chemical bonds to yield smaller molecular. The equipment for pyrolysis or thermal cracking is expensive for modest throughputs. In addition, though the products are chemically similar to petroleum-derived gasoline and diesel fuel, the removal of oxygen during thermal processing also eliminates any environmental benefits of using an oxygenated fuel (Ma and Haa, 1999). Transesterification (also called alcohosis) is the reaction of a fat or oil and an alcohocol (with or without catalyst) to form esters and a by-product, glycerol. A catalyst is usually used to improve the reaction rate and yield. As The reaction is reversible, excess alcohol is used to shift the equilibrium to the products side (Vyas *et al*., 2001). Transesterification consists of 3 stepwise with 2 intermediates formation of diglycerides (DG) and monoglycerides (MG). The 3 steps are shown in Figure 2.2.

$$\text{Triglyceride (TG)} + \text{ROH} \xrightleftharpoons{\text{Catalyst}} \text{Diglyceride (DG)} + \text{RCOOR}_1$$

$$\text{Diglyceride (DG)} + \text{ROH} \xrightleftharpoons{\text{Catalyst}} \text{Monoglyceride (MG)} + \text{RCOOR}_2$$

$$\text{Monoglyceride (MG)} + \text{ROH} \xrightleftharpoons{\text{Catalyst}} \text{Glycerol (GL)} + \text{RCOOR}_3$$

Figure 2.2 Three steps of transesterification reaction

In the transesterification of biodiesel, a triglyceride reacts with an alcohol in the presence of a strong acid or base, producing a mixture of fatty acids alkyl esters and glycerol. The overall process is a sequence of three consecutive and reversible reactions, in which di- and monoglycerides are formed as intermediates. The stoichiometric reaction requires 1 mole of a triglyceride and 3 mole of the alcohol. However, an excess of the alcohol is used to increase the yields of the alkyl esters and to allow its phase separation from the glycerol formed.

The mechanism of the base-catalyzed transesterification of vegetable oils is shown in Figure 2.3. The first step is the reaction of the base with the alcohol, producing an alkoxide and the protonated catalyst. The nucleophile attack of the alkoxide at the carbonyl group of the triglyceride generates a tetrahedral intermediate in second step, from which the alkyl ester and the corresponding anion of the diglyceride are formed in third step. The latter deprotonates the catalyst, thus regenerating the active species, which is now able to react with a second molecule of the alcohol, starting another catalytic cycle. Diglycerides and monoglycerides are converted by the same mechanism to a mixture of alkyl esters and glycerol (Schuchardt et al., 1998).

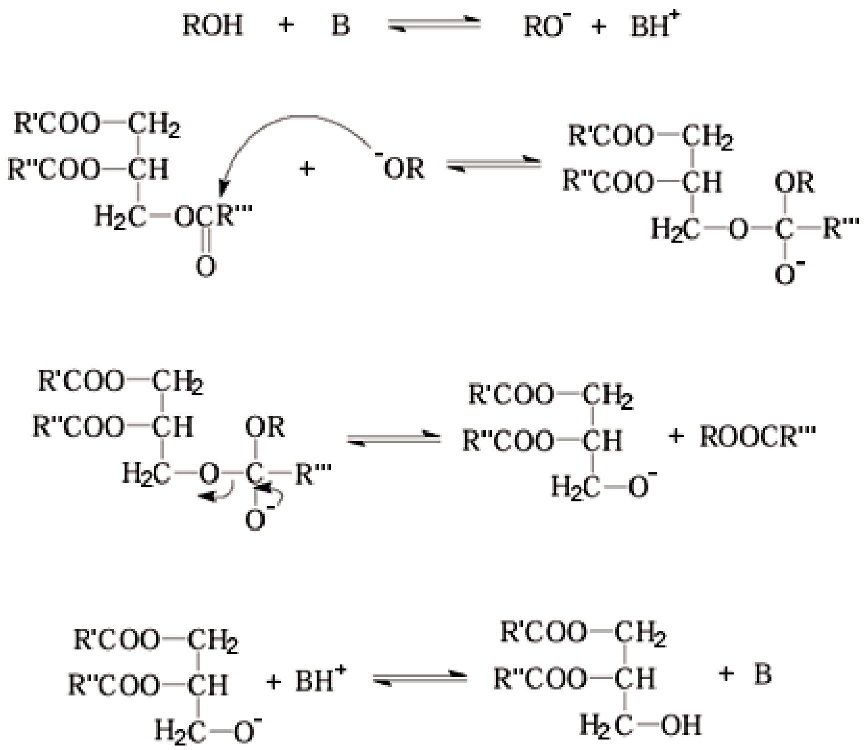


Figure 2.3 The mechanism of alkali catalyst transesterification (Schuchardt *et al.*, 1998)

Transesterification can be processed by either homogeneous or heterogeneous methods. Both methods have different advantages. Marchetti *et.al.* (2007) studied the possible methods for biodiesel production. For the homogeneous method, there are several kinds of catalyst such as alkali-catalyzed and acid-catalyzed. The advantages of alkali-catalyzed method are the fast reaction, high conversion and less corrosive than acid-catalyzed but there is a side-reaction; saponification. The problem of the soap formation can be solved by using acid-catalyzed because it does not have OH^- which is the reason of problem. However the acid-catalyzed is slow reaction, corrosive catalyst and high

condition
(e.g. pressure and temperature). For the heterogeneous method, there are several kinds of catalysts proposed. Lipase is one of heterogeneous catalysts. The advantages are the easy separation between catalyst and glycerin. Nevertheless, the disadvantages of lipase catalyst are slow reaction and expensive cost. Other catalysts in heterogeneous process are $Na/NaOH/g\text{-}Al_2O_3$, $ZrO_2\text{-}SiO_2$, $KOH/ZrO_2\text{-}SiO_2$, $Co_2O_3\text{-}SiO_2$, $Mo_2O_2\text{-}SiO_2$, $Na_2O\text{-}SiO_2$, La_2O_3-MCM-41 and MgO.

The kinetic model of biodiesel formation was proposed by many researches (Freedman, 1986; Noureddini and Zhu, 1997; Darnoko and Munir, 2000; Foon *et al*., 2004; Karmee *et al*., 2004; Vicente *et al*., 2005; Vicente *et al*., 2006). The general form of governing differential equations involving three steps is following:

$$\frac{d[TG]}{dt} = -k_1[TG][A] + k_2[DG][E] \quad (1)$$

$$\frac{d[DG]}{dt} = k_1[TG][A] - k_2[DG][E] - k_3[DG][A] + k_4[MG][E] \quad (2)$$

$$\frac{d[MG]}{dt} = k_3[DG][A] - k_4[MG][E] - k_5[MG][A] + k_6[GL][E] \quad (3)$$

$$\frac{d[GL]}{dt} = k_5[MG][A] - k_6[GL][E] \quad (4)$$

$$\frac{d[E]}{dt} = k_1[TG][A] - k_2[DG][E] + k_3[DG][A] - k_4[MG][E] + k_5[MG][A] - k_6[GL][E] \quad (5)$$

$$\frac{d[A]}{dt} = -\frac{d[E]}{dt} \quad (6)$$

where A and E are the alcohol and ester concentrations; respectively.

In addition, the activation energy and reaction rate shown in the Tables 2.3 and 2.4 were estimated by the Arrhenius equation.

$$k(T) = AT^n e^{-E/RT} \quad (7)$$

Table 2.3 The energy of activation and reaction rate constant at 65 °C using KOH (Vicente, 2005)

Reaction	Energy of Activation (J/mole)	Reaction rate constants (L/mol.min)
TG → DG	31,656.2	1.5
DG → TG	31,014.3	13.7
DG → MG	41,557.8	23
MG → DG	41,107.2	41.4
MG → GL	5,955.5	0.4

2.1.6 Benefits of biodiesel

There are many benefits of using biodiesel rather than the traditional petroleum diesel as follows:

2.1.6.1 Biodiesel can be used in the most diesel equipment with no or only minor modifications therefore it can directly replace the petroleum diesel, reducing the country's dependence on imported oil. (http://www.uoregon.edu/sskach/ Benefits.html)

2.1.6.2 Biodiesel can reduce global warming gas emission such as carbon dioxide (CO_2). CO_2 is taken up by the annual production of plants like soybeans and then released when vegetable oil based biodiesel is burned. This cycle does not add to the net CO_2 concentration in the air because the next crops will reuse the CO_2 in order to grow. In addition, biodiesel also reduce tailpipe particulate matter (PM), hydrocarbon (HC) and carbon monoxide (CO) emissions. Pure biodiesel (B100) contains 11% oxygen by weight. The presence of fuel oxygen allows the fuel to burn more completely therefore unburned fuel is insignificantly produced. (U.S. Department of Energy, 2006)

2.1.6.3 Biodiesel is nontoxic, biodegradable and suitable for sensitive environment. Pure biodiesel can breaks down four times faster than regular diesel, with most of a spill broken down only 28 days later. It is also less sulfur dioxide than petroleum based diesel, and and sulfur dioxide is the major component of acid rain. (http://www.terrabioenergy.com/benefits/index.php)

2.1.6.4 Biodiesel is safer to transport and handle, because it has a higher flash point (great than 150^{o}C) than traditional petroleum (77^{o}C), and is non-corrosive for human contact. (http://www.terrabioenergy.com/benefits/index.php)

2.1.6.5 Biodiesel is more lubricating than diesel fuel that result from the free fatty acids present, which reduces engine wear and extends engine life. (Zuhair, 2007)

2.1.7 Technology of Commercial Biodiesel Production (Barker, 2005)

Technology of commercial biodiesel production can be classified by 3 production processing, i.e. batch, on-line by transesterification and on-line by 2 stages (esterification and transesterification)

Batch Technology

Advantage is the low cost but the product quality may not be steady and a production capacity per round is not large

On-line Technology by Transesterification

A process with a steady product quality required an installation space less than the batch technology at equal capacity but at higher investment cost.

On - line Technology by 2 stages

Apply the esterification process at first stage and then transesterification process at a second stage. This technology is suitable with all raw materials, esp. oil with a high value of free fat acidic

2.2 Batch Processes (Barker, 2005)

Today many produces using continuous processes were originally produced using batch processes. The main reason behind the shift from batch process to continuous process was that the batch processes were labor-intensive and required skilled and experienced operation to produce batch products with consistency in quality. However, due to increasing demand for flexible and customer-driven production, batch processes finally slow equally the importance in manufacturing industries. Batch processes are economical for small-scale production as it requires few number of process equipment and intermediate storage is inexpensive. Batch processes are suitable for manufacturing of large number of products or special products due to flexibility in manufacturing process equipment.

Batch processes in practical have the following characteristics:

1. Batch processes deal with discrete quantities of raw materials or products.
2. Batch processes allow the tracking of these discrete quantities of materials or products.
3. Batch processes allow more than one type of product to be processed simultaneously, as long as the products are separated by the equipment layout.
4. Batch processes entail movement of discrete product from one processing area to the other.
5. Batch processes have recipes (or processing instructions) associated with each load of raw material to be processed into product.
6. Batch processes have more complex logic associated with processing than is found in continuous processes.

2.2.1 Classification of Batch Processes

Batch processes can be classified on the basis of two criteria:

- The quantity of output produced
- The structure of batch process plant

2.2.2 Batch manufacturing basics

One of the objectives of batch process automation is to manufacture a batch in optimum (minimum) time and to maximize the capacity utilization of the batch manufacturing facility. Investment in batch process automation can be justified by the returns resulting from the improvements in recipe scheduling which are easy to quantify. Various planning and scheduling techniques employing fairly simple to very complex algorithms are used for optimization of batch manufacturing processes in the industry. The planning and scheduling techniques have been developed and used for specific applications, depending on suitability to a particular batch process or an industry.

Batch numbering, tracking and reporting

- Batch tracking

The main objective of batch tracking is to provide another view of operations organized by specific raw materials, and intermediate and finished products. This is important capability as the problems that are not apparent in daily operating result can become visible when viewing the operation by product. To make product tracking possible, the transactions in each of the operational areas of receiving, production, inventory, transfers and shipments must be properly time-sequenced.

- Batch records/history

The batch plant must all data gather related to the production in each batch. This may be done in both hard copy from and electronic archives. Batch history requirements are more demanded in the pharmaceutical industry, where all batch events that occurred during production must be capture, including all procedural events such as batch start, hold, restart and complete, process alarms, operator changes, operator comments, recipe procedure execution events, material consumption, material production and trends related to key process variables.

2.2.3 Batch process management

Typically, production scheduling is tactical and deal with detailed timing of special manufacturing step, while campaign planning is more strategic and related to controlling costs over longer periods of time. Both concepts are characterized by extensive data needs, uncertainty, a large decision space (sequencing, timing, product assignment to unit, etc.) and the need for good, feasible solutions. Optimal plans and schedules, found through numerical modeling techniques, may not always be required to satisfy the real-world business needs even though they might be worth the effort. Depending on the sophistication of approach used to solve the problems, a feasible plan or schedule may be all that is considered necessary to meet the immediate business needs.

Production scheduling is the short-term look (less than a week to a mouth) at the requirements for each product to be made. The time scale should fit the needs of manufacturing. Decisions that must be made at this level are:

- How many batches required for each product
- Which equipment to use if multiple units are available
- Start and stop time of each batch piece of equipment (the run length)
- Allocation of resources to support the production of those batches (e.g. utilities, operation, raw materials, waste facilities)

Campaign planning, then is a medium-term look (week to moths) at a series of batches

2.2.4 Batch planning and scheduling

Batch scheduling has attracted wide attention of the scheduling research community over the past two decades. Every plant executes based on a production schedule .The schedule may come from a planning entity somewhere in the company and it is usually produced on regular schedule such as monthly, weekly or daily. The main objective of batch production planning and scheduling is to optimize capacity utilization of batch manufacturing facilities and fulfill customer orders within time.

There are several excellent papers of reviews and comparative studies on scheduling of chemical processes.

Floudas and Lin (2004) compared discrete and continuous-time approaches for scheduling of multiproduct/multipurpose batch and continuous processes. They examined slot-based and precedence-based models in sequential process scheduling, and compared various event-based models in network process scheduling.

Lee *et al*. (1992) provided polynomial-time algorithms for the problems of scheduling jobs with agreeable processing times and due dates on a single batch processing machine to minimize maximum tardiness and number of tardy jobs.

Li and Lee (1997) proved that the batch scheduling problem with agreeable release dates and deadlines is strongly NP-hard. They developed polynomial-time algorithms for the problems of minimizing maximum tardiness and number of tardy jobs when all the jobs have agreeable release dates, due dates and processing times.

2.2.4.1 Classification of batch scheduling problems

There are a great variety of aspects that need to be considered when developing scheduling models for batch processes. In order to provide a systematic characterization present first a general roadmap for classifying most relevant problem features.

2.2.4.2 Classification of optimization models for batch scheduling

Having presented the general features of typical batch scheduling problems introduces a roadmap that describes the main features of current optimization approaches. This section is of particular importance because alternative ways of addressing/formulating the same problem are described. These usually have a direct impact on computational performance, capabilities and limitations of the resulting optimization model.

2.3 Economic analysis (Seider *et.al,* 2010)

The economic analysis is one of the most important steps that must be determined in order to have a profitable plant investment which produces valuable products. Firstly, the capital investment, which is total cost that requires setting up the plant and other facilities, must be considered to build all aspects of the facility necessary for this plant. The capital investment composes of fixed-capital investment (FCI) and working capital (WC). The fixed-capital investment is the investment of manufacturing and facilities which can be divided into two sections as direct cost and indirect cost. The working capital is the amount of money which is required to operate plant in the first period. The sum of fixed-capital investment and working capital is total capital investment (TCI).

2.3.1 Estimation of capital investment cost (Towler, 2008)

2.3.1.1 Capital Investment Costs

The total capital investment (TCI) of the chemical plant is a one-time expense for design, construction, and startup of a new plant.

The capital required to supply manufacturing and plant facilities is called the *fixed-capital investment* (FCI) ,while that necessary for the operation of the plant is teamed the *working capital* (WC). The sum of the fixed-capital investment and the working capital is known as the total capital investment (TCI). It is the sum of fixed-capital portion may be further catagorized into *manufacturing fixed-capital investment*, also known as *direct cost*, and *nonmanufacturing fixed-capital investment*, also known as *indirect cost*.

Fixed-Capital Investment

Direct cost shows the capital necessary for the installed process equipment with all components that are required for complete process operation. Expenses for site preparation, piping, instrument, insulation, foundations, and auxiliary facilities are typical examples of costs included in the direct cost.

The capital spent for construction overhead and for all plant components that are not directly related to the process operation is designate the indirect cost. These plant components include the land; processing building, administrative and other office, warehouses, laboratories, transportation, shipping, and receiving facilities, utility and waste disposal facilities, shops and other permanent parts of the plant. The *construction overhead cost* includes field office and supervision expenses, home office expenses, engineering expenses, miscellaneous construction costs, contractor's fees, and contingencies. In some case construction overhead is proportioned between manufacturing and indirect cost.

Investment cost

1. Purchased equipment

All equipment listed on a complete flowsheet depends on many factor such as size, power of equipment moreover mass of catalyst that is used in operation.

2. Purchased-equipment installation

Total cost of installed equipment that listed on a complete flowsheet, structure support and equipment insulation and painting is called $C_{TBM.}$

3. Site preparation

Site preparation typically involves making land surveys, dewatering and drainage. Surface cleaning, rock blasting, excavation, grading, piling; and addition of fencing, roads, sidewalks, railroad sidings, sewer lines, fire protection facilities, and landscaping. Cost of site preparation and development, C_{site}, can be quite in substantial for exiting integrated complex, the cost only 5% of the total cost of the installed equipment.

4. Building (including services)

Process building; Substructure, superstructure, platforms, supports, stairways, ladders. access way, cranes, monorails, hoists and elevator.
Auxiliary building; administration and offices, medical or dispensary, cafeteria, garage, product warehouse, parts warehouse, guard and safety, fire station, change house, personal building, shipping office and platform, research laboratory and control laboratory.

Maintenance shop; Electric, piping, sheet metal, machine, welding, carpentry and instrument.

Building services; Plumbing, heating, ventilation, dust collection, air conditioning, building lighting, elevators, escalators, telephones, intercommunication systems, painting, sprinkler systems and fire alarm. Cost of building, $C_{building}$, is only 5% of the total cost of installed equipment.

5. Offsite

Offsite cost, $C_{offsite}$, are include to provide or upgrade offsite utility plants (steam, electricity, cooling water, process water, boiling feed water, refrigeration, inert gas, fuels, etc.) and related facilities for liquid waste disposal, solids waste diposal, off-gas treatment and wastewater treatment. It is about 5% of the total cost of installed equipment.

6. Land

The cost of land, C_{land}, is non-depreciable, since land rarely decreases in value, and in the absence of data can be taken as 2% of the depreciable capital, C_{TDC}.

7. Royalties

The license fee may be a one-time fee, in which case that fee is included in the capital investment as a one-time royalty or paid up license, C_{royal}. An initial royalty fee of 2% of C_{TDC}.

8. Startup

The cost of plant startup, $C_{startup}$, is typically estimated as 10% of $C_{TDC.}$ However, according to Feldman (1969), if the process and equipment are well known to skilled operators and the new process is not dependent on the operation of another plant, the startup cost may be as low as 2% of $C_{TDC.}$

Working Capital

The working capital for industrial plant consists of the total among of money investment.

1. Raw material and supplied carried in stock.
2. Finished product in stock and semi-finished products in the process of being manufactured.
3. Accounts receivable.
4. Cash kept on hand for monthly payment of operating expenses, such as salaries, wages, and raw material purchases.
5. Accounts payable
6. Taxes payable

2.3.1.2 Estimation of total product cost, C

The total annual production cost equals the sum of the cost of manufacture and general expenses,

$$C = COM + GE \tag{15}$$

Costs of Manufacturing, COM

All expenses directly connected with the manufacturing operation or the physical equipment of process plant is included in manufacturing cost. These expenses of manufacturing contain 3 major categories as (1) direct production costs; (2) fixed charges; and (3) plant overhead costs.

Direct production costs

1. Feed stock (raw materials)

In the chemical industry, one of the major costs in a production operation is for raw material used in the process that are directly consumed in making the final product. The amounts of raw materials that must be supplied per unit of time or per unit of product are determined form process material balances. One of the most important steps of the design process is to calculate accurate material balances for process.

2. Utility

The required types of utilities are established by the flowsheet conditions; their amount can sometimes be estimated in preliminary cost analyses from available information about similar operation. More often the utility requirements are determined from material and energy balances calculated for the process. A utility may be purchased at a predetermined rate from an outside source, or the service may be available within the company.

3. Labor-related operations, O

One of the most difficult annual cost to estimate is direct wages and benefits (DW&B) for operating a chemical plant. DW&B calculated from an hourly rate for the operators of a proposed plant. To estimate all labor-related operations, it is necessary to estimate the number of operators for the plant per shift and to account for four shift per daily. Estimates of the number of plant operators needed per shift are based on the type and arrangement of the equipment and the multiplicity of units.

4. Maintenance, M

A second category of labor-related costs is associated with the maintenance of a proposed plant. Processing equipment must be kept in acceptable working order, with repair and replacement of parts made as needed. Annual maintenance costs, M, are something greater than the cost of labor-related operations, O. The maintenance costs consist of maintenance wages and benefits (MW&B), which is estimated as 3.5% of C_{TDC}. Salaries and benefits for the engineers and supervisory personnel are estimated at 25% of MW&B. Material and services are estimated at 100% of MW&B, while maintenance overhead is estimated at 5% of MW&B.

Fixed changes

1. Depreciation, D

For use with approximate profitability measures, as applied here to the preliminary calculation of annual manufacturing cost, depreciation, D, is estimated as a constant percentage of the total depreciable capital, C_{TDC}. This type of depreciation is referred to as *straight-line (SL) depreciation*. The direct plant (on-site) depreciation is taken as 8% of (C_{TDC}-1.18$C_{offsite}$) ,while the allocated plant (offsite) depreciation is taken as 6% of the contribution of 1.18 $C_{offsite}$.

2. Property Taxes and Insurance

Annual property taxes are assessed by the local municipality as a percentage of the total depreciable capital, C_{TDC}, with a range from 1-3%. Property taxes are not related to federal income taxes levied by the Internal Revenue Service and considered below. Liability insurance costs depend on the pressure and temperature levels of plant operation and on weather flammable, explosive, or toxic chemical are involved. The property taxes and insurance may be estimated at 2% of C_{TDC}.

Plant overhead costs

Overhead expenses include the cost of providing the following service: cafeteria; employment and personnel; fire protection, inspection and safety; first aid and medical; industrial relation; janitorial; purchasing, receiving, and warehousing; automotive and other transportation; and recreation. Overhead costs are divided into four categories: general plant overhead can be estimated at 7.1% of M&O-SW&B, provision for the service of the mechanical department can be estimated at 2.4% M&O-SW&B and for the employee relations department can be estimated at 5.9% M&O-SW&B, as well as business services can be estimated at 7.4% M&O-SW&B. M&O-SW&B can be estimated the combined salary, aged, and benefits for maintenance and labor-related operation.

General Expenses, GE

General expenses referred to activities that are conducted by the central operations of a company, perhaps at the corporate headquarters, and are financed from profits made by the company from their operating plant. General expenses comprise five categories: selling (or transfer) expense can be estimated at 2% of sales, research (direct 4.8% of sales and allocated 0.5% of sales), administrative expense can be estimated at 2% of sales, and management incentive compensation expense can be estimated at 1.25% of sales.

2.3.1.3 Total incomes

Total incomes came from product, by product and something which can sell.

2.3.1.4 Gross earnings cost

Amount of gross earnings cost depends on amount of gross earning for entire company and income tax regulations.

$$\text{Gross earning} = \text{total income} - \text{total product cost} \tag{16}$$

2.3.2 Profitability Analysis

Profitability analysis is used to consider the economic feasibility after the cost estimation analysis is done. This is also used to select an appropriate type of investment which is less risky, more reliable and worthwhile.

The first reason for the investors is try to make the most income in their investment. On the one hand, it should be at least more than the profit from saving money in a bank. There are 2 ways to calculate the profitability. First one is the method that does not consider the time value of money. Therefore, the inflation rate (or discount rate) will be neglected. The other method will consider the time value of money. Consequently, the inflation rate is included this calculation as well.

2.3.2.1 Depreciation, D

The depreciation must be determined due to it is related to the profitability calculation. There are 5 methods used to estimate the depreciation rate include the straight line method (SL), the declining-balance method (DB), the double declining-balance method (DDB), the sum-of-the-years digits method (SYD), Accelerated Cost Recovery System (ACRS) and a Modified Accelerated Cost Recovery System (MACRS). The ACRS and MACRS methods combine aspects of the DB or DDB methods with the SL method. The most effective method is the last method as it is more accurate. Accordingly, the Modified Accelerated Cost Recovery System (MACRS) is used in this project. Percentage of total depreciable capital, C_{TDC} for MACRS is shown in Table 2.4.

2.3.2.2 Payback Period

Payback period is the length of time required for cumulative incoming returns to equal the cumulative costs of an investment (e.g. purchase of computer software or hardware, training expenses, or new product development), usually measured in years.

Other things being equal, the investment with the shorter payback period is considered the better investment. The shorter payback period is preferred because:

- The investment costs are recovered sooner and are available again for further use.
- A shorter payback period is viewed as less risky. It is usually assumed that the longer the payback period, the more uncertain are the positive returns. For this reason, payback period is often used as a measure of risk, or a risk-related criterion that must be met before funds are spent. A company might decide, for instance, to undertake no major investments or expenditures that have a payback period over, say, 3 years.

Table 2.4 MACRS Tax-Basis Depreciation

Year	**10 Year**
1	10.00
2	18.00
3	14.40
4	11.52
5	9.22
6	7.37
7	6.55
8	6.55
9	6.56
10	6.55
11	3.28
	100.00

2.4 Literature reviews

Chakton has performed the simulations of biodiesel process from Palm oil using reactive distillation to compare with conventional process both homogenous and heterogeneous catalyst via ASPEN PLUS, ASPEN DYNAMICS program and estimated unavailable component by GaussViewW and GAUSSIAN 03W program.
For the homogeneous transesterification process used the rate constants of Noureddini and Zhu (1997) while the heterogeneous process used the reaction rate of Petchtabtim (2008).

The simulation of the conventional process was followed by Zhou's ideal (Zhou, 2006). Two gravitational units were applied to the separation section. The process of conventional process with homogenous catalyst is shown in Figure 2.4. Homogenous conventional process started with mixing of methanol 130 kg/h with sodium hydroxide 12 kg/h to produce sodium methoxide while 1,000 kg/h of palm oil which were heated to 60 ˚C and sent to the reactor. Reactor temperature was maintained to 60 ˚C all the time. The reactor effluent was sent to a distillation in order to separate and recycle excess methanol. Biodiesel and glycerin from the bottom column were sent to purification section which glycerin was separated from biodiesel in the first gravitational unit. Higher density glycerin was sent from bottom liquid and lower density biodiesel containing some catalyst was mixed with water 900 kg/h and sent to the second gravitational unit to separate biodiesel and water-sodium hydroxide mixture.

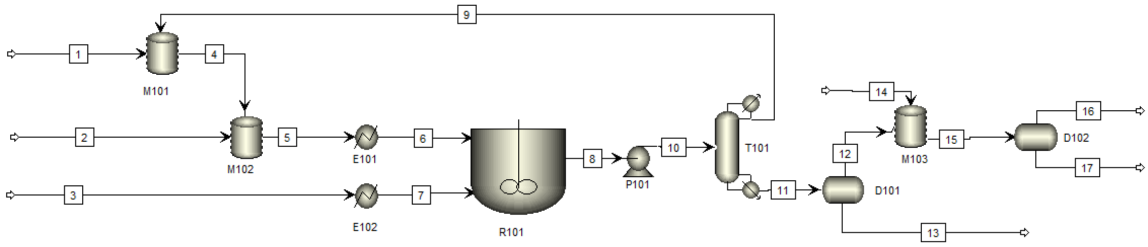


Figure 2.4 The process of conventional process with homogenous catalyst.

Sadude has studied the simulation and design of biodiesel production in terms of time schedule and cost estimation. The research was simulated with feedstock capacity of 100,000; 300,000 and 500,000 kg/cycle. Reactor, setting and mixing tanks were batch operating units meanwhile evaporator and centrifuge were continuous operating units. The feedstock of biodiesel production plant was stearin; methanol was main reactant for transesterification and potassium hydroxide was used as catalyst. This work consisted of 2 section. The first is fixed flow rate and second is fixed time. The results were shown the fixed flow rate pattern of operation gave the higher yield of biodiesel than that of fixed time one but fixed flow rate pattern yield higher total cost that than of fixed time pattern. For fixed time pattern, increasing amount of equipment and reducing size of equipment increased yield of biodiesel, but it was not be too much. On the other hand, for fixed flow rate, increasing amount of equipment and reducing size of equipment increased yield of biodiesel. For cost estimation of fixed time pattern, increasing amount of equipment and reducing size of equipment reduced total cost. However, the fixed flow rate pattern, increasing amount of equipment and reducing size of equipment increased total cost. Therefore, the chosen biodiesel production process should have low cost meanwhile it produced high amount of biodiesel. The ratio of total cost per yield of biodiesel was the lowest, that plant should be selected. Consequently, the biodiesel production of plant D at capacity 300,000 kg of stearin in fixed time pattern of operation was suitable for investment.

Table 2.5 Summery literature reviews

Literature	Catalyst		Process		Technical process		Method		Batch	Gly*
	Homo	Hetero	Esterification	Tranesterification	Conventional	Reactive Dist.	Experiment	Simulation		
Giovanni et al, 2010	/			/	/			/		
Petchtabtim, 2008		/		/	/		/			
Anton *et al*, 2008		/	/			/		/		
Limniyakul, 2007	/			/		/		/		
Huayang *et al*, 2007				/	/		/			
Chayasoontorn, 2007	/			/	/	/		/		
Alex *et al*, 2006	/	/		/	/			/		
Bhatia *et al*, 2006		/	/			/	/	/		/
Bournay *et al*, 2005	/	/		/			/			
Zhang *et al*, 2003	/			/	/			/		/
Noureddini and zhu, 1997	/			/	/		/			/
This work	/	/		/	/	/		/	/	/

*Process with glycerol purification section

CHAPTER 3 METHODOLOGY

This work relates to the simulation of biodiesel production. The process composes of a lot of equipment. However, the main equipment are a mixing tank, a reactor, a coalescer, an evaporator, a washing tank and a centrifuge. In the glycerol purification section, a reactor, filter and distillation were added in the process. The procedure of this work is presented in this chapter.

The study simulation of biodiesel production from palm oil consists of four parts: a steady state simulation, a batch simulation and an economic analysis as shown in Figure 3.1.

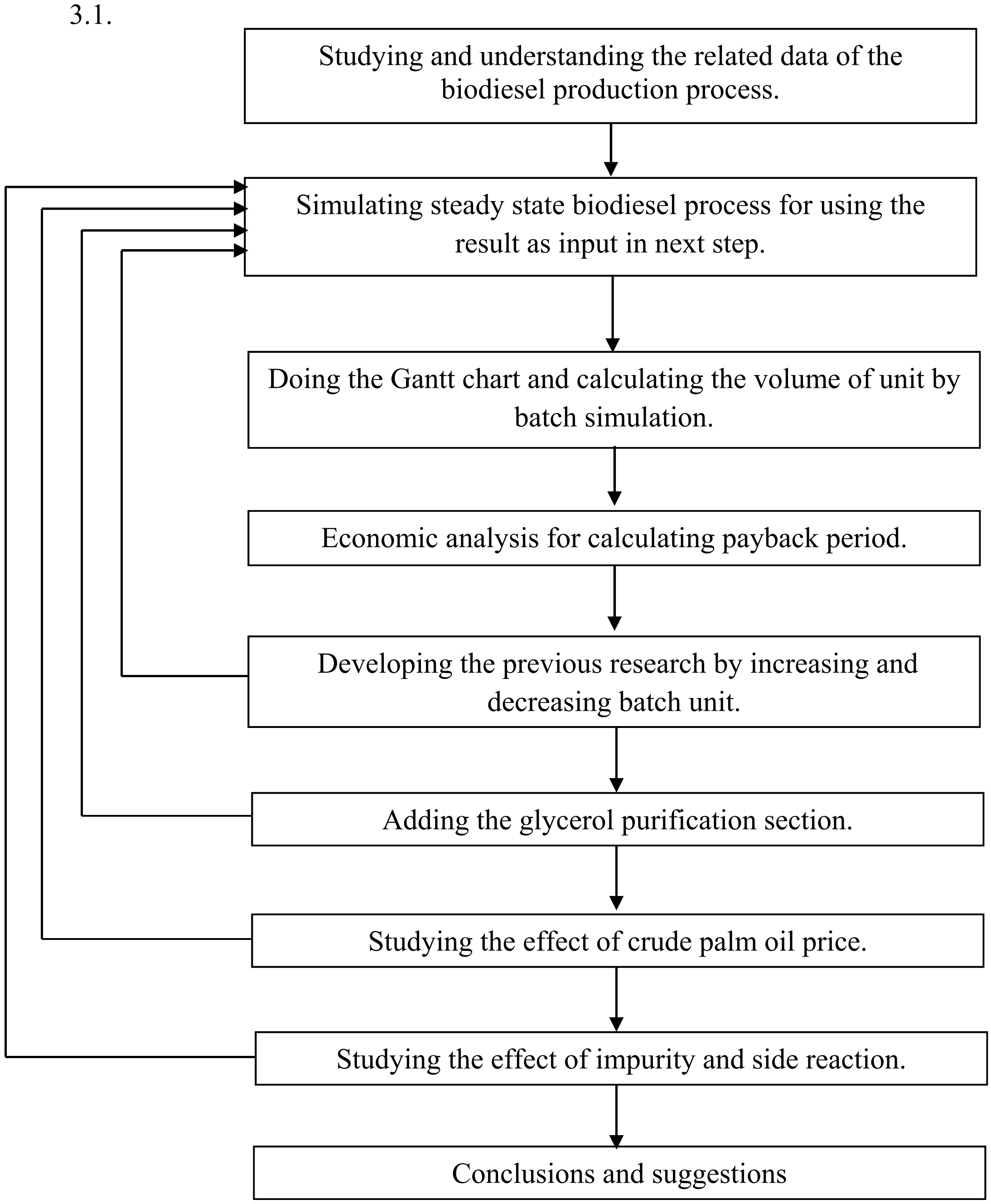


Figure 3.1 The methodology.

3.1 Studying and understanding the related data

This work starts by gathering the biodiesel and batch scheduling from the literatures. Standard biodiesel specification is used for the simulation. The biodiesel is normally produced by the transesterification method using homogenous catalyst, KOH, under conventional distillation. Then, the resident time of equipment is assumed from real plant and equipment size is calculated from the batch simulation. After that, the economic analysis is studied in order to propose the suitable process.

3.2 Simulation of a steady state biodiesel process

The part of the steady state is simulated to use to result as input in the batch simulation. The process flow sheet and data from laboratory of the previous research were simulated again in this step via ASPEN PLUS. The details of the biodiesel plant are clearly shown about the technique of biodiesel plant such as thermodynamic properties and operating condition in next chapter. After the steady state is complete, the batch simulation can be further performance.

The design Batch Scheduling of biodiesel production is carried out by Aspen Batch Plus version 7. The production processes consists of 3 main parts which are the batch unit such as mixing of methanol and KOH catalyst, reaction, settling glycerol and biodiesel and washing with water. For performing design and optimize batch scheduling, the standard eleven steps was followed the details were described as follow:

1. Start Batch Plus
2. Create a new project
3. Set up pure component data
4. Set up predefined mixture data
5. Set up equipment data
6. Set up reaction data
7. Select default units of measure
8. Create a new process
9. Create a new step
10. Simulate the Batch
11. Analyst the Results

3.3 Gantt chart and calculating the volume of unit by batch simulation

Some results from the steady state simulation were used in the batch process in order to calculate mass in and out of equipment. In addition to mass balance, the resident time and size of equipment were previously in step one. These were required to do the Gantt chart and to calculate the liquid volume size of each batch unit.

3.4 Economic analysis

After the Gantt chart is complete, batch cycle time and liquid volume size were calculated. Equipment cost can be calculated because equipment size can calculated from the volume size of each unit. The batch cycle time was used to calculate labor costs. Payback period is also an important factor as this can inform the suitability for investment.

3.5 Development and improvement of batch process

The previous research simulated the biodiesel process and consists of three mixing tanks, two reactors and one washing tank. In this section, the number of mixing tanks was varied to between 1 to 5. The number of reactors was varied to between 1 to 3 and washing tank were varied to between 1 to 2. Then, steady state, batch simulation and economic analysis were done again to calculate payback period.

3.6 Glycerol purification section

The previous research did not have the purification section which made off spec glycerol so this model cannot sell the glycerol. Although glycerol can sell, the investment cost and operating cost when this section is added is higher. Because of previous reason, payback period is considered again.

3.7 The effect of crude palm oil price

This section is studied the effect of crude palm oil price; the main raw material of the biodiesel production. The result from this section will cover the possibility of process depend upon palm oil price.

3.8 Studying the effect of impurity and side reaction

This section is studied to know the effect of impurity and side reactions. This research concerned only beta-carotene that affected the colour of glycerol and free fatty acids that is the substrate of side reaction, saponification. The result from this section can tell how much impurity in feed stock, plant should be shutdown.

3.9 Conclusions and suggestions

Finally, the results will be analyzed, concluded, and suggestions given for future work in this area.

CHAPTER 4 RESULT AND DISCUSSION

4.1 Simulating the base case

4.1.1 Steady state simulation

The previous research studied a process flow diagram is shown in Figure 4.1. There were consisted of 3 mixers, 2 reactors and 1 washing tank. Since this process has simulated by another feature of simulation so ASPEN PLUS has been held for this occasion. As mentioned to Figure 4.2, the thermodynamic method UNIFAC, ASPEN PLUS function, was selected for this duty and estimated Tri-palmitic, Tri- linoleic and Tri-oleic which are present in the crude palm oil. In order to produce 200,000 liter/day, the sufficient flow rate of crude palm oil, KOH and methanol had to be 8700, 90 and 1900 kg/h; respectively. Feed stock of palm oil with weight fraction of triolein, tripalmitic and trilinoleic are 0.4139, 0.4738 and 0.1123; respectively. The feeds were in the conditions at room temperature ($25^{O}C$) and 1 atm. Methoxide, which produced from KOH and methanol reacted in mixing tanks for 30 minutes, was used as a catalytic for this transesterification process. Then, the palm oil and methoxide were heated up to $65^{O}C$ before being sent to the reactors. The residence time for each reactor was 1.5 hours. Over the reactor circumstance, the rigorous continuous stirred tank reactor model (CSTR) was performed. The liquid phase reaction was isothermally reacted at 1 atm and $65^{O}C$. The kinetic parameters of transesterification reaction steps were previously mentioned in the chapter 2. The reactor effluent was transferred to a settling tank unit in order to separate the two liquid phases with biodiesel in the upper phase and glycerol in the lower phase. The separating process took 45 minutes for each settling tank. Both phases were transferred to an evaporator unit to recover and to increase the purity of methanol. Glycerol was produced in this step but due to its specification range so caused to worthless. Pure water stream, 4500 kg/h, was added-in the first washing tank to remove KOH. After that, the effluent from the washing tank was transferred to the the centrifugal unit to separate waste water and biodiesel. Then biodiesel was transferred to the washing tank and centrifugal unit again. In the final step, a high humidity biodiesel was transferred to an evaporator in order to reduce the moisture content to be in the standard level.

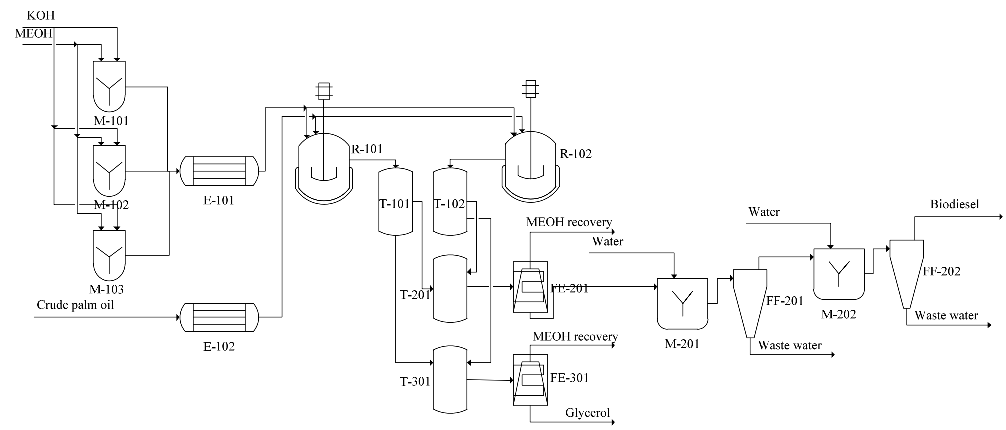


Figure 4.1 The base case process flow diagram of biodiesel production.

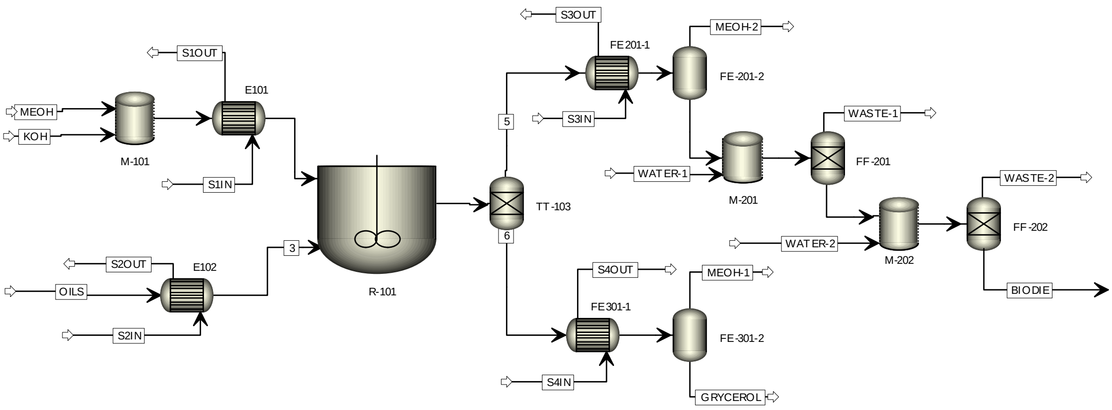


Figure 4.2 Process flow diagram of biodiesel via ASPEN PLUS.

4.1.2 Batch simulation

The process modeling and simulation system will perform the mass balance of each unit in the process. The term scheduling is used for actual assignment of resources to production tasks and sequencing and timing of these operations. Gantt chart displays the sequence and operating time of each unit. Schedule view consists of three parts. Top part showed time in operation each equipment and cycle time in system. Left part present the unit operation and bar chart represent the operating time of unit in line. Red colour is the first batch, blue is the second and green is the third batch. Three batches were computed by the same procedure. Starting with charging palm oil, KOH and methanol from the truck into the storage tank at the same time. Feed flow rate of palm oil, KOH and MEOH are 22,700, 50 and 85 cubic meter per hour; respectively. After that, one third of KOH and methanol were transferred to Mixer1 with 5 kg/h and 85 kg/h with the mixing time at 30 min. Repeating the same procedure in mixer 2 and 3. The methoxide were transferred to reactors 1 and 2. The resident times for each reactors are 1.5 hours. Next, the 2 reactor effluents were transferred to 2 settling tanks, 200 kg/h. This unit uses for 45 min. The upper layer, crude biodiesel, was transferred to

TT-201-1 and TT-201-2 with flow rate 1000 kg/h while the lower layer (glycerol) was transferred to TT-301 with flow rate 200 kg/h. The crude biodiesel was fed to evaporator FE-201 which is the continuous unit. Condensed methanol was kept in TT-104 and transferred this biodiesel to the washing tank M201. Water was added-in to remove catalyst for 60 minutes. Centrifuge the contents of M201 in centrifuge FF201-1 to separate waste water. After finishing second washing step, the biodiesel was obtained as the product. The crude glycerol from settling tank was transferred to evaporator for recovering methanol. The second batch can run before the first batch finish because related units are available. Resident time of second washing tank is the key controlled of the total reaction time since it needs the longest time to perform. The period of batch production which does not the scheduling is 3 times greater than the scheduling batch process. Normal batch process has to wait finished product from final the unit and then new raw material is added in first unit the scheduling batch process does not wait the product finished. Figure 4.3 is the Gantt chart of the base case with scheduling. The biodiesel production factory in 3D animation as illustrated in Figure 4.4, can be useful for stepping through a plan chronologically schedule, the recipe description, and the material balance for biodiesel production.

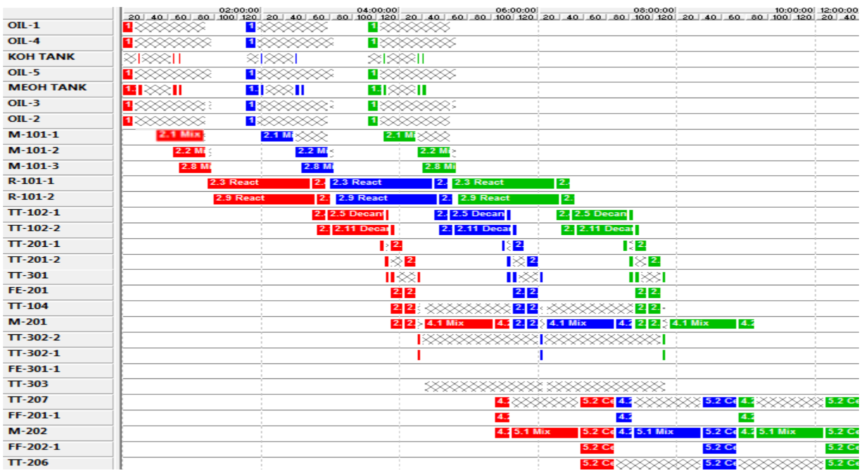


Figure 4.3 Batch scheduling of base case.

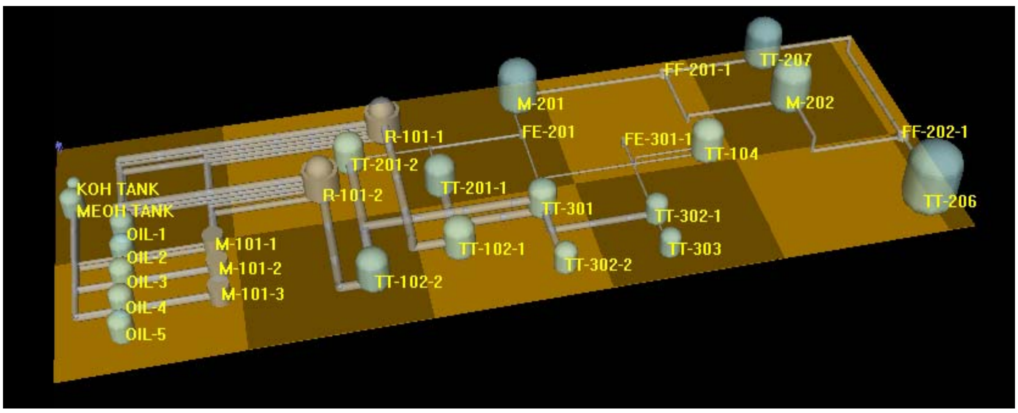
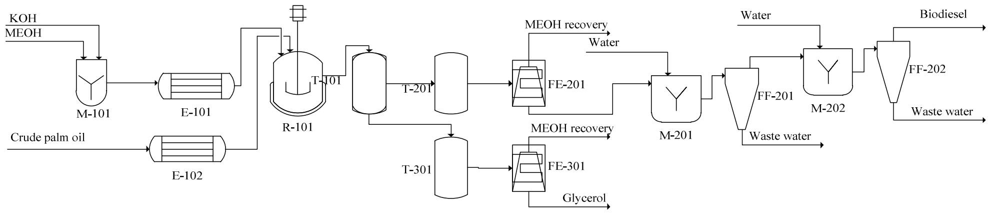


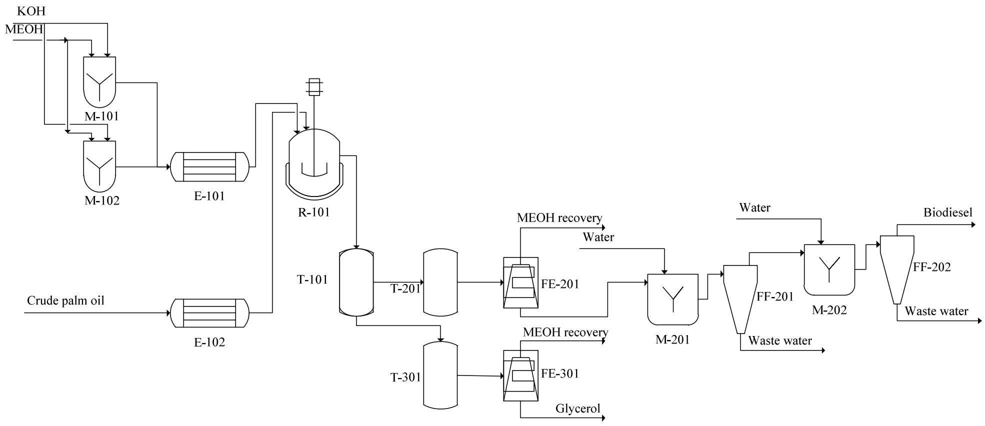
Figure 4.4 3D animation of base case

4.2 Various number of batch unit for the batch unit.

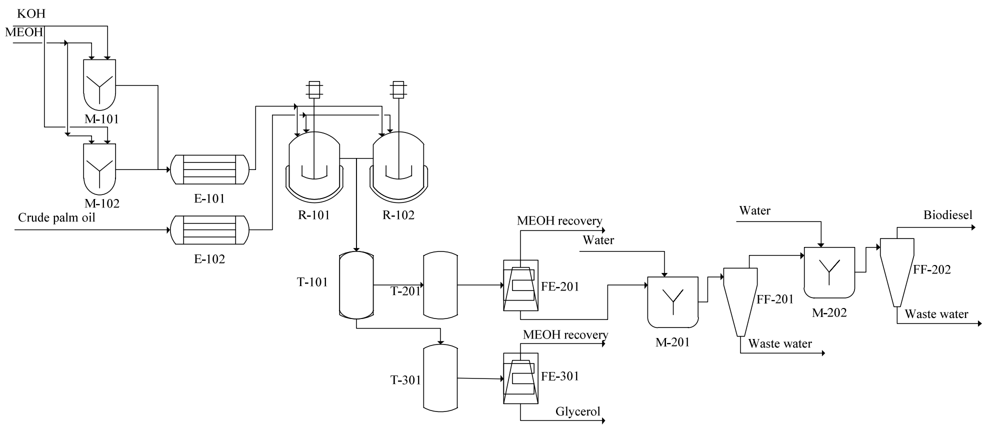
The base case was consisted of 3 mixers, 2 reactors and 1 washing tank. In this study, we varied number of the varying unit by the batch units from 1 to 5 mixers, 1 to 3 reactors and 1 to 2 washing tank. All of those units were splited in parallel layout in order to decrease the batch time. There were a total of thirty new models. Some of these processes are illustrated in Figure 4.5.



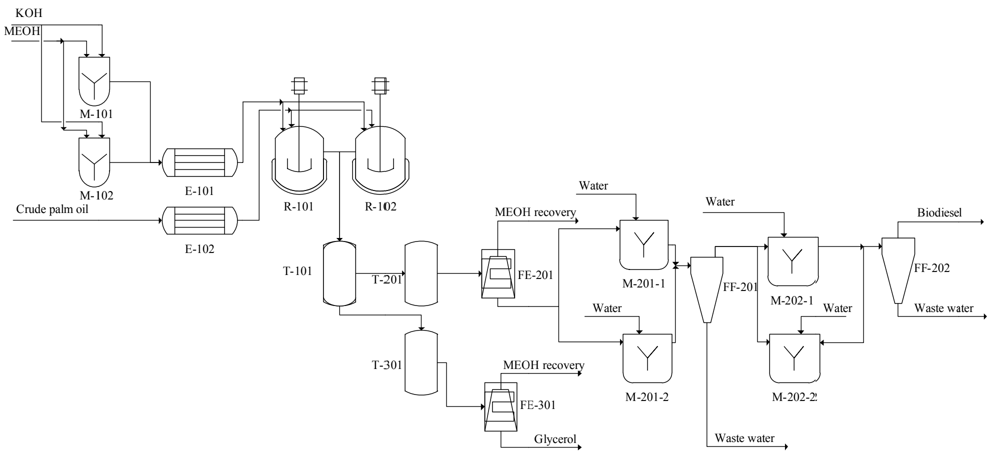
(a)



(b)



(c)



(d)

Figure 4.5 The process flow diagram of biodiesel production
(a) The process with 1 mixer, 1 reactor and 1 washing tank
(b) The process with 2 mixers, 1 reactor and 1 washing tank
(c) The process with 2 mixers, 2 reactors and 1 washing tank
(d) The process with 2 mixers, 2 reactors and 2 washing tanks

Gantt chart of new process is shown in APPENDIX D. The operating time of all process is not different because the resident time of each unit is the same. After all simulations and economics were done, the results were shown in Figures 4.6 through 4.8. Figure 4.6 was investment cost. Figure 4.7 was batch time and Figure 4.8 was the payback period that consists of figures depending on interest rate 10%.

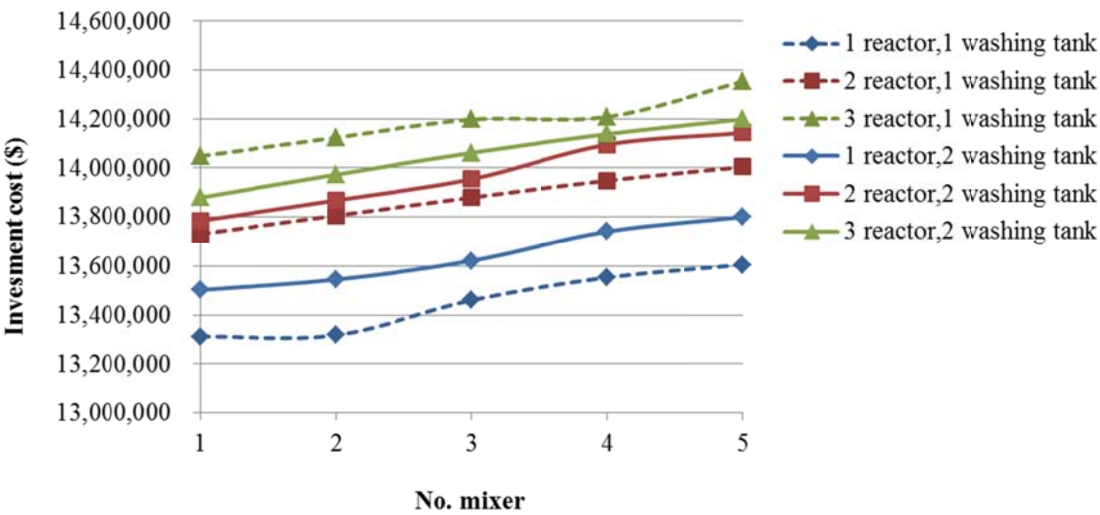


Figure 4.6 Investment cost of new biodiesel processes

Figure 4.6 shows the number of reactors and washing tanks are varied. Although using more units, both of reactors and washing tank would require higher investment cost, the number of reactors rather than the number of mixers had a greater impact on batch time. However, the investment cost is calculated from the Chemical Design Book (Seider, 2010). With based on the expression, cost of reactor and washing tank depend on volume of liquid in the tank. Obviously, the only one unit caused the small number of the cost comparing to the system with more than one unit. In the operating basis, if only one reactor is used in the process, the big size of the reactor would be required and totally become to the high cost due to its complexibility of mechanical fabricates.

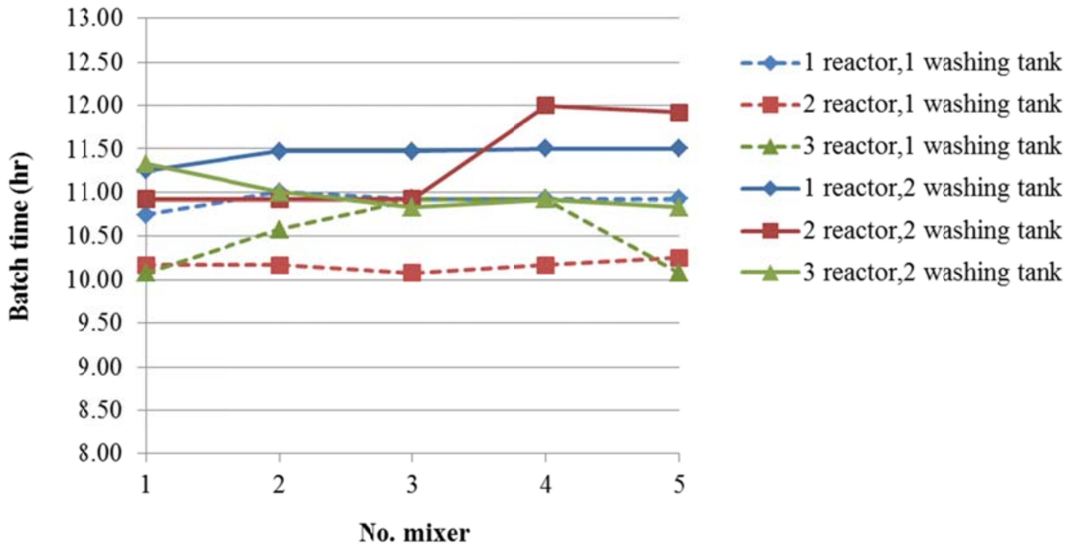


Figure 4.7 Batch time of new biodiesel processes

Prefer to Figure 4.7, the process with one reactor took longer payback period comparing to another processes, two and three reactors against both of 1 and 2 washing tanks. It noticed that the system consists of 2 washing tanks, the batch time parameter will be higher than 1 washing tank. The reason for phenomenon is caused by the critical time which will control the total time of the process. For 2 washing tanks, there has critical time at the tank while 1 washing tank has the critical time at reactor.

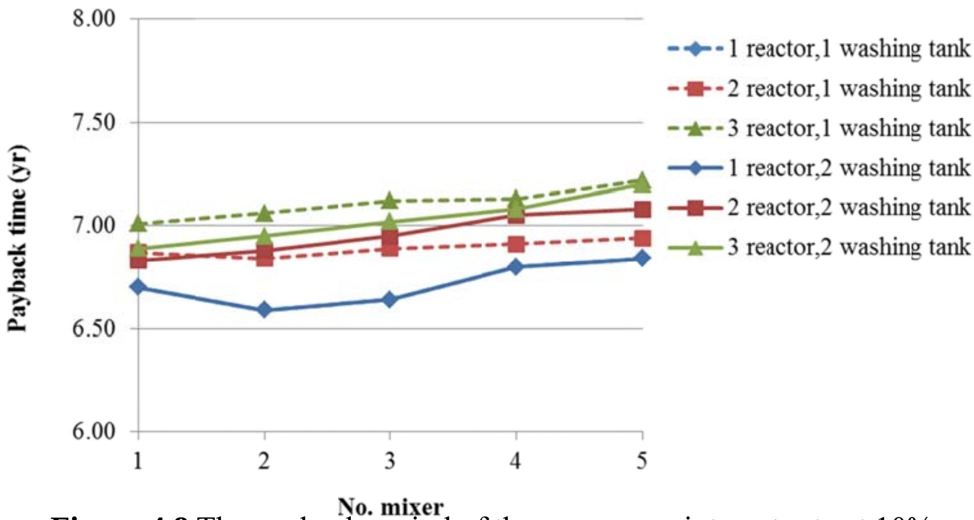


Figure 4.8 The payback period of the process on interest rate at 10%

According to Figure 4.8 the trend of the payback period shows that the equipment, the biggest time consumption when adding more equipment to the process. Approaching to more reactors then more taken time of payback period, there are the same in terms of added more mixing tank and also washing tank. Regarding the above mention that is the results from payback period has direct-modified to the investment cost, since the constant benefit even we added more units.

The comparison between those processes which had one washing tank and two washing tanks found that the number of washing tanks did not affect to the payback period when the interest rate was low level but it became to have influence when the interest rate was higher. For example, at 18.5% interest rate has affected to payback period time much more than 15% and 10% interest rate; consequence.

4.3 The glycerol purification section

This section was introduced to increase the purity of crude glycerol in order to sell the glycerol as another product. The glycerol purification section composed of a reactor, filter and distillation column as shown in Figure 4.9.

The crude glycerol from the settling tank effluent was transferred to a neutralize reactor together with H_3PO_4. This reaction produced solid K_3PO_4 and water at 1 atm, 25 ^{O}C. K_3PO_4 was separated by filler while the unreacted glycerol and water were transferred to distillation column at stage 3. The distillation column had 6 stages and operated at 0.4 atm using 1 mass reflux ratio and 0.47 molar bottoms to feed ratio. Investment cost and payback period are shown in Figure 4.10 and 4.11.

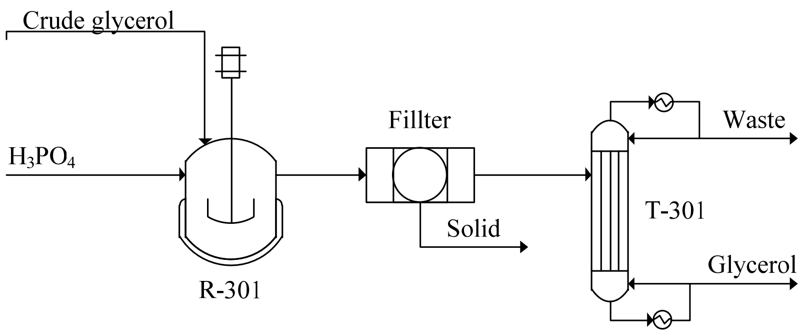


Figure 4.9 Process flow diagram of glycerol purification section

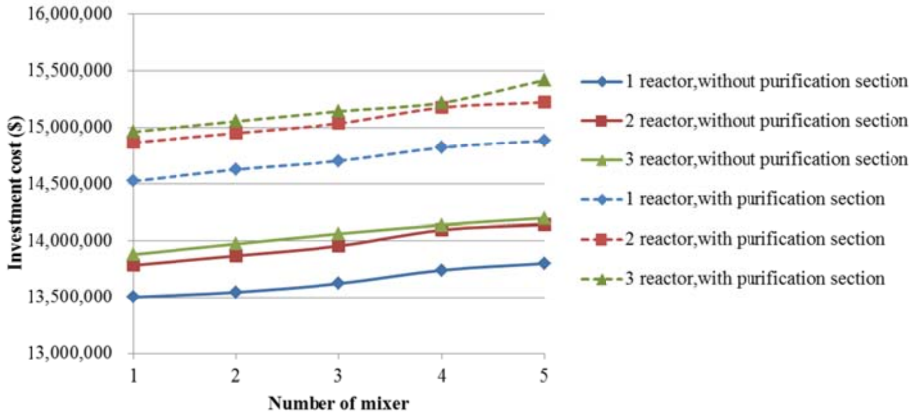
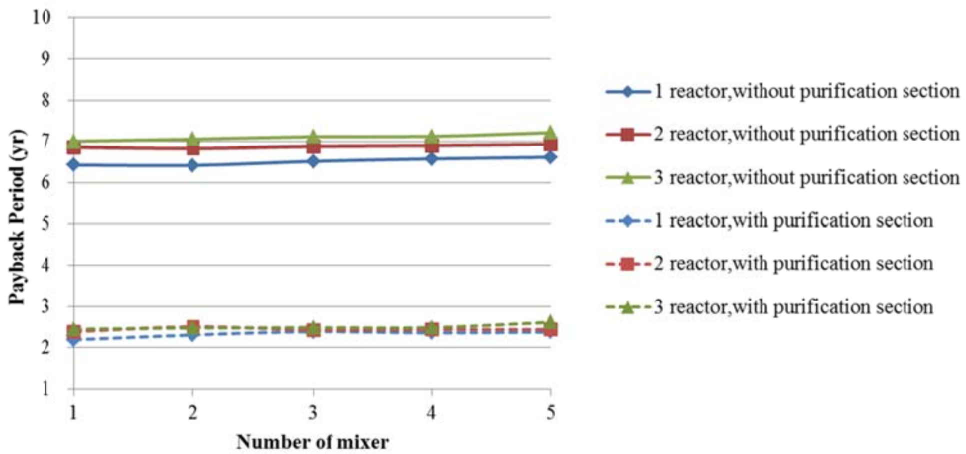
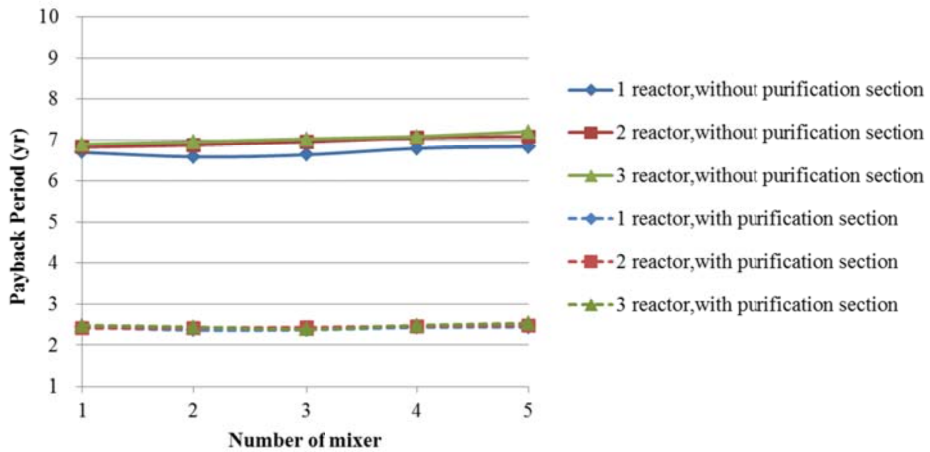


Figure 4.10 Comparison of investment cost between process with and without glycerol purification section of process composed 1 washing tank



(a)



(b)

Figure 4.11 Comparison of the payback period between process with and without glycerol purification section (a) with 1 tank (b) with 2 washing tanks

The process which used a purification section required more investment cost (around 1,000,000 USD will be added up), However, it can decrease the payback period by 5 years at low interest rate (10% interest rate), 7 years at the middle interest rate (15% interest rate) and 10 years at high interest rate (18.5% interest rate). The shorter in number of payback period is advantaged from the opportunity to serve/sell the high analysis grade of glycerol which is the by product with 99.5% purity. The number of mixers, washing tanks and interest rate slightly affected to the number payback period in the process which included the glycerol purification section. The process with the purification section was more suitable than the process without this unit.

4.4 The effect of crude palm oil price

As the previous section, the investment cost and the payback period were considered to determine the proper model, however, the palm oil price effect was also important. Palm oil price directly affected to the operating cost and the payback period. If the palm oil price was increased, the operating cost and the payback period were concordantly increased. The operating cost and the payback period of all models were nearly the same trend, e.g. , the operating cost and the payback period of the process which composes 3 mixers and 2 washing tank were demonstrated in Figure 4.12 and Figure 4.13; respectively.

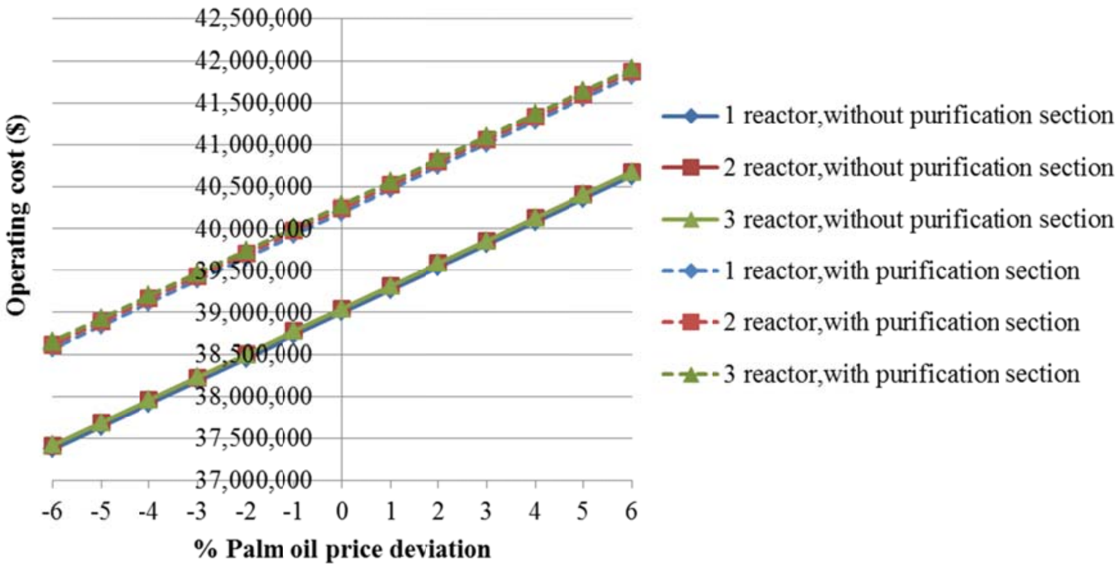


Figure 4.12 The operating cost of process which composes 3 mixers and 2 washing tanks

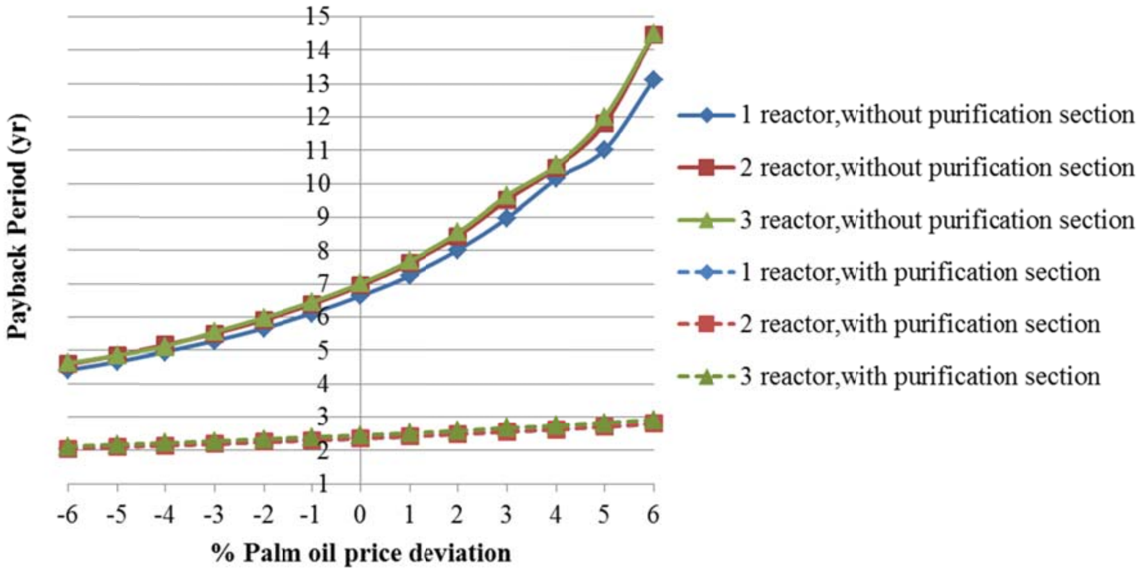
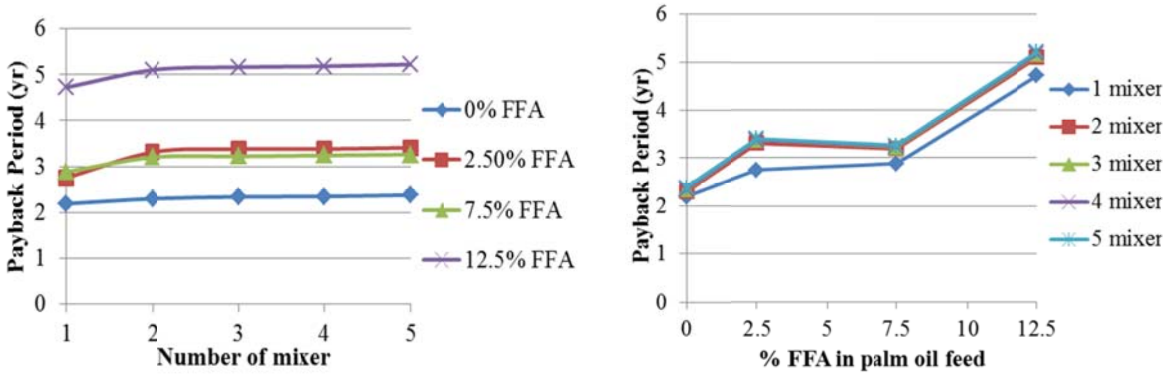


Figure 4.13 The payback period of process of 3 mixers and 2 washing tanks

When the payback period was fixed at 10 years, the process without purification section can support only 5% deviation of palm oil price while the process with purification section can support 27% deviation. All of this caused from the sensitive of palm oil price, when the palm oil price has raise up will directly affect to only the process without glycerol purification unit, since the process with unit will get the analysis glycerol 99.5% as the by product which can be sold for maintaining the capital cost. Therefore, the glycerol purification section should be added to the biodiesel process.

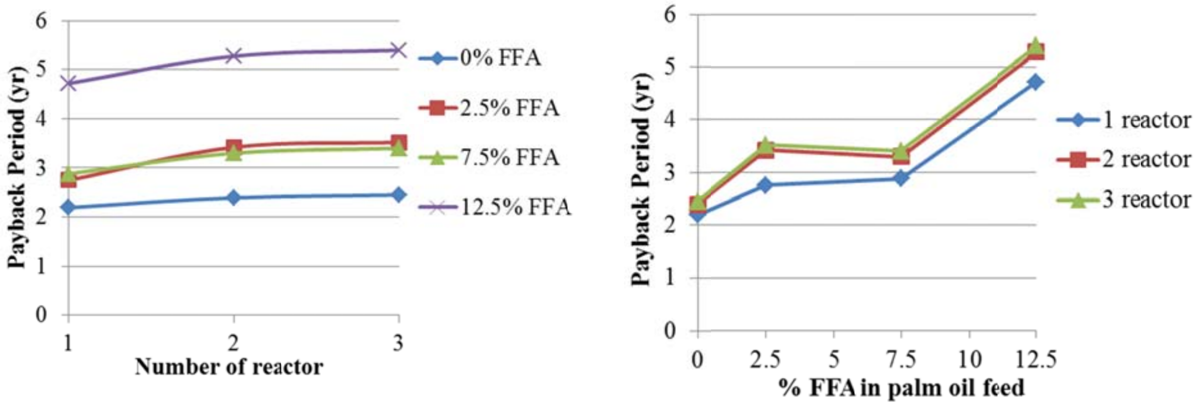
4.5 Studying the effect of impurity and side reaction

In this section, the processes with glycerol purification section were studied. Beta-carotene and free fatty acid, palmitic acid, were the important impurities. Beta carotene affected to glycerol physical colour. Free fatty acid is a reactant of saponification reaction which was side reaction. The composition of fatty acid was varied at 0%, 2.5%, 7.5% and 12.5% by weight with content beta carotene at 2.5% by weight. Figure 4.14 shows that the number of mixer did not affect to the payback period when changing amount of free fatty acid. Figure 4.15 shows that the process which had 2 reactors or 3 reactors took longer time than 1 reactor to get the same benefit. Both Figures, 4.14 and 4.15, demonstrated the payback period depends on the amount of free fatty acid in palm oil feed. If the palm oil feed contained free fatty acid between 2.5% to 8.5%, the payback period was rarely affected but if the amount of free fatty acid was more than 8.5%, it strongly affected to the payback time.



(a) (b)

Figure 4.14 The payback period of process consisted 1 reactor and 1 washing tank



(a) (b)

Figure 4.14 The payback period of process consisted 3 mixers and 1 washing tank

CHAPTER 5 CONCLUSIONS AND RECOMMENDATIONS

5.1 Conclusions

In this work, the process simulations consisted of two parts which were steady state and batch simulations. ASPEN PLUS simulator was used to simulate the steady state part while ASPEN BATCH was used in batch part. Then, economic analysis was calculated in order to determine appropriated model. Many criteria were considered before investing such as net present value, internal rate of return, profitability index and payback period. Payback period was used as criteria in this work.

There were two main processes which were the biodiesel process with and without glycerol purification section. Both main processes were divided into 30 models which were varied number of mixer, number of reactor and number of washing tank. The mixer of Olein, Linolein and Palmitin were used in this studied to present the Palm oil. Gantt chart, 3D Animation of schedule, recipe description and liquid volume of batch units were calculated via batch simulator. After the simulations were done, payback period was an important factor which used in this work in order to determine proper model. For the comparison between the two main processes, with and without the glycerol purification section; the payback period of the process with glycerol purification section was lower than the process without purification section because the process gets a high grade glycerol which was high value. In addition, the process with purification section can absorbed higher palm oil price more than process without purification section when changing the price. The process which had 1 mixer is suitable for investment because number of mixer did not affect payback period but the more unit, the more investment cost. 1 or 2 washing tank rarely affect the payback period when lower interest rate but fairly affect when higher interest rate. 1 washing tank was appropriate because it was shorter payback time and spent lower investment.

In practical, the palm oil feed did not consist of Olein, Linolein and Palmitin but also had impurity. This work concerned only beta carotene which affects the colour of glycerol and free fatty acid, palmitic acid. Palmitic acid is a reactant of saponification reaction which was side reaction of transesterification. The payback period depend on the amount of free fatty acid in palm oil feed. If the palm oil feed was 2.5% to 7.5% free fatty acid, it rarely affects but if the amount of free fatty acid was more than 7.5%, it strongly affects payback time.

In conclusion, the biodiesel process with glycerol purification section consists of 1 mixer, 1 reactor and 1 washing tank is optimum for investment in biodiesel production.

5.2 Recommendations

5.2.1 This work estimated the properties of feed component such as Tri-palmitic, Tri-linoleic, Tri-oleic and etc., via ASPEN PLUS. If the actual properties of these were input in simulation, the result will be more reliable.

5.2.2 In economic part, this work calculated following the text book which it does not suitable in Thailand.

5.2.3 After the steady state was done, it should optimize the operation condition.

5.2.4 In the end of the batch simulation should design the controller.

REFERENCES

Agreda, L.R. Partin and W.H. Heise. 1990. High-purity methyl acetate via reactive distillation. **Chemical Engineering Progress.** 86: 40–46.

Alex H. West, Dusko Posarac, Naoko Ellis, 2006. Assessment of four biodiesel production process using HYSYS.Plant. **Bioresource Technology.** 99:6587-6601.

Anton A. Kiss, Alexander C. Dimian, Gadi R., 2008. Biodiesel by Catalytic Reactive Distillation Powered by Metal Oxides. **Energy & Fuels**. 22: 598-604.

Barker, J. Rawtani and S. mackay. 2005. **Practical Batch Process Management**. Elsevier, Netherlands.

Bell, M. Blinick , E. Creber, G. Greer, E. Ho and M. Knight. 2007. **Biodiesel.** Queen's University. 90.

Benefits of biodiesel. Available source: http://www.terrabioenergy.com/benefits/index.php, November 25, 2009 .

Bhatia, A.L. Ahmad, A.R. Mohamed and S.Y. Chin. 2006. Production of isopropyl palmitate in a catalytic distillation column: Experimental studies. **Chemical Engineering Science.** 60: 7436-7447.

Bournay L.,Cassanave D., Delfort B., Hillion G.. Chodorge J.A., 2005. New heterogeneous process for bio diesel production: A way to improve the quality and the value of the crude glycerin produced by biodiesel plants. **Catalyst Today**. 106: 190-192.

Chakton K., 2009. **Simultion of Biodiesel Production from Palm Oil by Conventional and the Reactive Distillation Methods.** Master of Engineering (Chemical Engineering), Department of Chemical Engineering, Kasetsart University.

Chayasoontorn K., 2007. **Plantwide Control for Continuous Biodiesel Process**. Master of Engineering (Chemical Engineering), Department of Chemical Engineering, King Mongkut's University of Technology Thonburi.

Darnoko, D. and C. Munir. 2000. Kinetics of Palm Oil Transesterification in a Batch Reactor. **JAOCS.** 77: 1263-1267.

Demirbas, A. 2008. Biodiesel fuels from vegetable oils via catalytic and non-catalytic supercritical alcohol transesterifications and other methods: a survey. **Energy Conversion & Management.** 44: 2093-2109

Foon, C. Y. May, M. Ahnagan and C. C. Hock. 2004. Kinetics Study on Transesterification of Palm Oil. **Journal of Oil Palm Research.** 16: 19-29.

Freeman, B. 1986. Transesterification kinetics of soybean oil. **JAOCS.** 63: 1375-1380.

Fukuda,H., A. Kondo and H. Noda. 2001.Boidiesel fule production by transesterification of oil. J. **Bioscience Bioeng.** 9 : 405-416.

Giovanni D.N., Matteo M., Marco P., Giulio S., 2010. Bioenergy II: Modeling and Multi-Objective Optimization of Different Biodiesel Production Processes. **Chemical Reactor Engineering**. Vol.8. Article A16.

Halek, F., A. Kavousi and M. Banifatemi. 2009. Biodiesel as an alternative fuel for diesel engines. **Engineering and Technology**. 57: 460-462.

Huayang He, Shiyao Sun, Tao Wang, Shenlin Zhu, 2007.Transesterification Kinetics of soybean Oil for Production of Biodiesel in Supercritical Methanol. **Amer Oil Chem Soc.** 84: 399-404.

Karmee, S.K., P. Mahesh, R. Ravi and A. Chadha. 2004. Kinetic Study of the Base-Catalyzed Transesterification of Monoglycerides from Pongmia Oil. **JAOCS.** 81: 425-430.

Limniyakul W., 2007. **Application of reactive Distillation for Biodiesael Production Enhancement.** Master of Engineering (Chemical Engineering), Department of Chemical Engineering, Kasetsart University, pp.24-64.

Ma, F. and M.A. Hanna. 1999. Biodiesel production: a review. **Bioresource Technol.** 70: 1-15.

Marchetti, V.U. Miguel and A.F. Errazu. 2007. Possible methods for biodiesel production. **Renewable & Sustainable Energy Reviews.** 11: 1300-1311.

Meher, D. Sagar and S.N. Naik. 2004. Technical aspects of biodiesel production by Transesterification-a review. **Renewable and Sustainable Energy Reviews.** 10:1-21.

Noureddini, H. and D. Zhu. 1997. Kinetics of Transesterification of Soybean Oil. **JAOCS.** 74: 1457-1463.

Petchtabtim, K., 2008 **Study of kinetic rate of heterogeneous catalysts for biodiesel production.** M.S. Thesis, Kasetsart University.

Schuchardt ,R. Sercheli and R.M. Vargas. 1998. Transesterification of vegetable oil: a review. **J.Braz.Chem. Soc**. 9: 199-210.

Seider, W.D. 2010. **Product and Process Design Principles**. Third edition. John Wiley & Sons, New Jersey.

Towler,G. and Sinnott, R., 2008. **Principles, Practice and Economics of Plant and Process Design.** Forth edition. Elsevier Inc.

U.S. Department of Energy. 2006. **Biodiesel Handling and Use Guidelines**. 3rd ed. U.S. Department of Commerce, Springfield.

Vicente, G., M. Martinez., J. Aracil and A. Esteban. 2005. Kinetics of Sunflower Oil Methanolysis. **Energy & Fuels.** 44: 5447-5454.

Vicente, G., M. Martinez and J. Aracil. 2006. Kinetics of Brassica carinata Oil Methanolysis. **Energy & Fuels.** 20: 1722-1726.

Vyas, A.P., J.L. Verma and N. Subrahmanyam. 2010. A review on FAME production Process. **Fuel** 89: 1-9.

Zhang, Y., Dube M.A., McLean D.D., Kates M., 2003. Biodiesel production from waste cooking oil: 1. Process design and technological assessment. **Bioresourse technology**. 89:1-16,

Zhou, H., Lu, H. and Liang, B., 2006, " Solubility of Multicomponent System in the Biodiesel Production by Transesterification of Jatropha curcas L. Oil with Methanol", **J. Chem. Eng. Data**, Vol.51, pp. 1130-1135.

Zuhair, S.A. 2007. Production of biodiesel: possibilities and challenges. **Biofuels Bioprod**. Bioref. 1: 57-66

APPENDIX

APPENDIX A

Summary Process Simulation

There are many processes were simulated. Table A.1 shows step to vary number of unit operation in biodiesel process. When glycerol purification section was added or changing the palm oil price or changing the percentage of impurity, Number of unit were varied following this table again.

Table A.1 Number of main unit in biodiesel process.

Process number	Number of unit operation		
	Mixer	Reactor	Washing tank
1	1	1	1
2	1	2	1
3	1	3	1
4	2	1	1
5	2	2	1
6	2	3	1
7	3	1	1
8	3	2	1
9	3	3	1
10	4	1	1
11	4	2	1
12	4	3	1
13	5	1	1
14	5	2	1
15	5	3	1
16	1	1	2
17	1	2	2
18	1	3	2
19	2	1	2
20	2	2	2
21	2	3	2
22	3	1	2
23	3	2	2
24	3	3	2
25	4	1	2
26	4	2	2
27	4	3	2
28	5	1	2
29	5	2	2
30	5	3	2

APPENDIX B

Economic Analysis

After the simulation has done, it has to do the economic analysis before investment. In this work, every process must do economic and calculate the payback period. The process which had 5 mixer, 3 reactor and 2 washing tank with glycerol purification section at no impurity was used be an example to describe economic calculation.

B.1 Equipment Cost

Mixers

Mixer Cost ($) = Tank Cost ($) + Motor Cost

Tank Cost ($) = 5700 + 700(volume, $m^3)^{0.7}$

Motor Cost ($) =1.46exp(4.2432+(2.03251(ln(Power,HP))) - (0.03595(ln(Power,HP))^2)

Table B.1 The cost of mixers in process

Unit ID	Size Utilized (L)	Tank Cost ($)	Cost of motor ($)	Total Cost ($)
M-101-1	3,994.01	7,545.37	948.488	8,493.86
M-101-2	3,994.01	7,545.37	948.488	8,493.86
M-101-3	3,994.01	7,545.37	948.488	8,493.86
M-101-4	3,994.01	7,545.37	948.488	8,493.86
M-101-5	3,994.01	7,545.37	948.488	8,493.86
M-201-1	42,463.30	15,354.01	5,968.31	21,322.3
M-201-2	42,463.30	15,354.01	5,968.31	21,322.3
M-202-1	51,389.58	16,733.43	6,803.26	23,536.7
M-202-2	51,389.58	16,733.43	6,803.26	23,536.7

Reactors

Reactor Cost ($) = Tank Cost ($) + Motor Cost($)

Tank Cost ($) = 14000 + 15400(volume, $m^3)^{0.7}$

Motor Cost ($) =1.46exp(4.2432+(1.03251(ln(Power,HP))) - (0.03595(ln(Power,HP))^2)

Table B.2 The cost of reactors in process

Unit ID	Size Utilized (L)	Tank Cost ($)	Cost of motor ($)	Total Cost ($)
R-101-1	30,001.09	180,539.27	4,671	18,5210
R-101-2	30,001.09	180,539.27	4,671	18,5210
R-101-3	30,001.09	180,539.27	4,671	18,5210
R-301	41,071.00	221,489.16	5,832	22,7321

Evaporator

Evaporator Cost ($) = 17000 + 13500(Heat transfer area, $m^2)^{0.6}$

Table B.3 The cost of evaporator in process

Unit ID	Area (m^2)	Cost ($)
FE-201	2,532	1,504,175

Centrifuges

Centrifuge Cost ($) = 63000 + 260000(Diameter, m)$^{0.8}$

Table B. 4 The cost of centrifuges in process

Unit ID	Diameter (m)	Cost ($)
FF-201	3	671,645
FF-202	3	645,880

Exchangers

Exchanger Cost ($) = 1030(Area, m^2)$^{0.6}$

Table B.5 The cost of exchangers in process

Unit ID	Area (m^2)	Cost ($)
E-101	2.32	1,704
E-102	2.01	3,166
Condenser	6.50	3,166
Reboiler	26.01	7,276

Pumps

Pump Cost ($) = 3300+48(Shaft work, kW)

Table B.6 The cost of pumps in process

Flow rate (m^3/h)	Shaft Work (kW)	Cost ($)	No.	Total Cost ($)
50	0.9	3,342	1	3,342
85	1.53	3,379	2	6,760
22700	408.26	81,591	5	407,960
5	0.09	3,302	1	3,303
5250	94.42	15,397	5	76,988
200	3.6	2,390	9	21,517
1000	17.99	5,456	3	16,368

Distillation Column

Distillation Cost ($) = Vessel Cost ($) + Drum Cost ($) + Tray Cost ($)

Vessel and Drum Cost ($) $= [1.67(0.959+0.041P-8.3x10^{-6}P^2)]x10^z$

z $= (3.17+0.2D+0.5\log_{10}L+0.21\log_{10}L^2)$

D = Diameter, m ; 0.3 m< D <4.0 m

L = Height, m ; L/D< 20

P = Absolute pressure, bar

Tray Cost ($) $= (187+20D+61.5D^2)N$

D = Vessel diameter, m

N = Number of tray

Table B.7 The total cost of distillation column in process

Unit ID	Amount	Vessel ($)	Drum ($)	Tray ($)	Costs ($)/unit	Total Cost($)
T-103	1	23,512	3,470	1,540	28,523	28,523
	Vessel					
	Diameter(m)	Height(m)	Absolute Pressure(bar)		Z	Cost($)
	0.91	7.32	1.03		4.15	23,512
	Tray					
	Diameter(m)	Number of tray	Cost($)			
	0.91	6	1,540			
	Drum					
	Diameter(m)	Height(m)	Absolute Pressure(bar)		Z	Cost($)
	0.91	0.91	1.03		3.32	3,470.91

Storage Tanks

Tank Cost ($) = 5700+700(Volume,m^3)$^{0.7}$

Table B.8 The total cost of storage tank in process

Unit ID	Volume size (L)	Cost ($)
OIL-1	14,007	10,142
OIL-2	14,007	10,142
OIL-3	14,007	10,142
OIL-4	14,007	10,142
OIL-5	14,007	10,142
KOH TANK	720	6,256
MEOH TANK	19,250	11,249
TT-102-1	30,001	13,270
TT-102-2	30,001	13,270
TT-102-3	30,001	13,270
TT-104	40,102	14,975
TM-201	18,450	11,086
TT-201-1	25,663	12,486
TT-201-2	25,663	12,486
TT-201-3	25,663	12,486
TT-206	200,000	34,264
TT-207	104,035	23,777
TT-301	13,015	9,919
TT-303	11,511	9,572

B.2 Cost Estimation

Table B.9 Total Capital Investment (TCI) of biodiesel production process

	Factor	Cost ($)
Total Equipment cost		4,698,418
Equipment erection	0.4	1,879,367
Piping	0.7	3,288,892
Instrumentation	0.2	939,684
Buildings	0.15	704,763
Offsite	0.05	234,921
Site	0.05	234,921
Total Physical plant cost		11,980,965
Startup	0.02	239,619
Contractor' fee	0.05	599,048
Contigency	0.1	1,198,097
Summation		2,036,764
Fixed capital		14,017,729
Working Capital	0.1	1,401,773
Total investment cost		15,419,502

Table B.10 The cost of raw material

Raw material	Flow rate (kg/yr)	Price ($/kg)	Cost ($/yr)
Palm oil	69,600,000	0.39	27,144,000
Methanol	15,200,000	0.4	6,080,000
H_3PO_4	431,179	0.8	344,943
KOH	720,000	0.98	705,600
Water	56,000,000	0.001	56,000
Total raw material cost			34,330,543

Table B.11 Direct Operating Labor Requirement for biodiesel plant

Equipment	Operators/unit/shift	Unit	Labor
Mixer	0.1	9	0.9
Reactor	0.5	4	2
Tower	0.5	3	1.5
Centrifuge	0.5	2	1
Pump	0	26	0
Heat Exchanger	0.1	2	0.2
Cooling towers	1	1	1
water treatment plants	2	1	2
Summary (man)			12.83

Time to operated (hr)	0
Salary ($/hr)	1500
Total operator cost ($/yr)	432,000

Table B.12 The cost of utility base on operating hour 8,000 hour/year

Utility	Type	Required	Unit	Cost / unit	Unit	Cost $/y
E-101	LP	250	kg/h	0.0023	$/kg	4,600
E-102	LP	250	kg/h	0.0023	$/kg	4,600
Reactor	Cooling	635	kW	4.43	$/GJ	18,282
Mixer	Electricity	900	HP	0.06	$/kWh	320,500
FE-201	LP	650	kg/h	0.0023	$/kg	11,960
T-301	HP	500	kg/h	0.0066	$/kg	26,400
FE-401	LP	300	kg/h	0.0023	$/kg	5,520
M-201	Electricity	225	HP	0.06	$/kWh	80,303
M-202	Electricity	1,000	HP	0.06	$/kWh	57,936
FF-201	Electricity	75	kW	0.06	$/kWh	5,358
FF-202	Electricity	40	kW	0.06	$/kWh	19,266
Waste treatment		7	m^3/h	0.056	$\$/m^3$	3,136
Total utility cost						450,000

Table B.13 The price of product

Product	Unit/day	Unit/year	Price($/unit)	Unit	Cost ($/yr)
Biodiesel	200,000	48,000,000	0.8426	Liter	40,444,800
Glycerol >99%	19,851	4,764,303	1.70	Liter	8,094,550
Methanol recycle	804	6,429,840	0.4	kg	2,571,936
Total revenue					51,111,286

Maintenance (M)

Wages and benefits (MW&B)

fluids handling process	0.03	of C_{TDC}	359,429
Salaries and benefits	0.25	of MW&B	89,857
Material and serviced	1.00	of MW&B	359,429
Maintenance overhead	0.05	of MW&B	17,971
Total Maintenance			826,687

Operating Overhead			
General plant overhead	0.071	of M&O-SW&B	58,695
Mechanical department services	0.024	of M&O-SW&B	19,840
Employee relations department	0.059	of M&O-SW&B	48,775
Business services	0.074	of M&O-SW&B	61,175
Total operating overhead			188,485
Property taxes and insurance	0.02	of CTDC	239,619
Depreciation (D)			
Direct plant	0.08	of (CTDC-1.18Calloc)	936,301
Allocated plant	0.06	of 1.18Calloc	14,095
Total D			950,396
COST OF MANUFACTURE (COM)			36,466,764
General Expenses			
Salling expense	0.01	of sales	511,113
Direct research	0.048	of sales	2,453,342
Allocated research	0.005	of sales	255,556
Administrative expense	0.02	of sales	1,022,226
Management incentive compensation	0.0125	of sales	638,891
Total Gen Expent	**GE**		4,881,128
Total Product Cost	com+ge		1,347,892
Profit			9,763,394

Table B.14 Depreciation of Biodiesel plant by MACRS method

Year	% of C_{TDC}	Depreciation ($/y)	Taxes Saved ($/yr)
2011	-	-	-
2012	10.00	1,541,950.21	462,585.06
2013	18.00	2,775,510.37	832,653.11
2014	14.40	2,220,408.30	666,122.49
2015	11.52	1,776,326.64	532,897.99
2016	9.22	1,421,678.09	426,503.43

Year	% of C_{TDC}	Depreciation ($/y)	Taxes Saved ($/yr)
2017	7.37	1,136,417.30	340,925.19
2018	6.55	1,009,977.38	302,993.21
2019	6.55	1,009,977.38	302,993.21
2020	6.56	1,011,519.33	303,455.80
2021	6.55	1,009,977.38	302,993.21
2022	3.28	505,759.66	151,727.90
2023	3.28	505,759.66	151,727.90
2024	3.28	505,759.66	151,727.90
2025	3.28	505,759.66	151,727.90
2026	3.28	505,759.66	151,727.90
2027	3.28	505,759.66	151,727.90
2028	3.28	505,759.66	151,727.90

Table B.15 Calculation of cash flows (Nominal interest rate = 15% and Income tax rate = 30%)

Year	Investment		D	$C_{excl.\ Dep.}$	S	Net Earnings	Discounted Cash Flow	Cash Flow (PV)	Cum. PV
	fC_{TDC}	C_{WC}							
2011	15,427,348	2,037,800	-	-	-	-	-17,465,148	-17,465,148	-17,465,148
2012	-	-	1,542,735	40,246,465	51,111,286	6,525,461	8,068,196	7,015,822	-10,449,326
2013	-	-	2,776,923	40,246,465	51,111,286	5,661,529	8,438,452	6,380,682	-4,068,644
2014	-	-	2,221,538	40,246,465	51,111,286	6,050,299	8,271,837	5,438,867	1,370,223
2015	-	-	1,777,231	40,246,465	51,111,286	6,361,314	8,138,544	4,653,239	6,023,462
2016	-	-	1,422,402	40,246,465	51,111,286	6,609,694	8,032,096	3,993,371	10,016,833
2017	-	-	1,136,996	40,246,465	51,111,286	6,809,478	7,946,474	3,435,480	13,452,313
2018	-	-	1,010,491	40,246,465	51,111,286	6,898,031	7,908,523	2,973,107	16,425,420
2019	-	-	1,010,491	40,246,465	51,111,286	6,898,031	7,908,523	2,585,310	19,010,730
2020	-	-	1,012,034	40,246,465	51,111,286	6,896,951	7,908,986	2,248,227	21,258,957
2021	-	-	1,010,491	40,246,465	51,111,286	6,898,031	7,908,523	1,954,866	23,213,823
2022	-	2,037,800	506,017	40,246,465	51,111,286	7,251,163	9,794,981	2,105,365	25,319,188
2023	-	-	506,017	40,246,465	51,111,286	7,251,163	7,757,180	1,449,872	26,769,060
2024	-	-	506,017	40,246,465	51,111,286	7,251,163	7,757,180	1,260,759	28,029,819
2025	-	-	506,017	40,246,465	51,111,286	7,251,163	7,757,180	1,096,312	29,126,131
2026	-	-	506,017	40,246,465	51,111,286	7,251,163	7,757,180	953,315	30,079,445
2027	-	-	506,017	40,246,465	51,111,286	7,251,163	7,757,180	828,969	30,908,415
2028	-	-	506,017	40,246,465	51,111,286	7,251,163	7,757,180	720,843	31,629,258

$C_{excl.\ Dep}$ = Total product cost – Depreciation; from Table 2.4, Chapter 2.

Net earnings = (S- $C_{excl.\ Dep}$-D) x (1-Income tax rate)

Annual cash flow = (net earnings + D) - fC_{TDC}- C_{WC}

APPENDIX C

Summary result

C.1 Process without glycerol purification section, No impurity

Table C.1 The investment of process with 1 washing tank in us dollar

No. mixer	No. reactor		
	1	2	3
1	13,311,570	13,731,014	14,048,952
2	13,317,226	13,806,524	14,124,463
3	13,460,344	13,879,795	14,197,733
4	13,553,345	13,948,903	14,208,320
5	13,605,574	14,004,403	14,354,229

Table C.2 The investment of process with 2 washing tanks in us dollar

No. mixer	No. reactor		
	1	2	3
1	13,311,570	13,731,014	14,048,952
2	13,317,226	13,806,524	14,124,463
3	13,460,344	13,879,795	14,197,733
4	13,553,345	13,948,903	14,208,320
5	13,605,574	14,004,403	14,354,229

Table C.3 The batch time of process with 1 washing tank

No. mixer	No. reactor		
	1	2	3
1	10.75	10.17	10.08
2	11.00	10.17	10.58
3	10.92	10.08	10.92
4	10.92	10.17	10.92
5	10.92	10.25	10.08

Table C.4 The batch time of process with 2 washing tanks

No. mixer	No. reactor		
	1	2	3
1	11.25	10.92	11.33
2	11.47	10.92	11.00
3	11.47	10.92	10.83
4	11.50	12.00	10.92
5	11.50	11.92	10.83

Table C.5 The Operating cost of process with 1 washing tank

No. mixer	No. reactor		
	1	2	3
1	38,994,538	39,074,815	39,045,091
2	38,987,486	39,040,309	39,046,585
3	38,997,475	39,041,755	39,048,032
4	38,999,311	39,043,119	39,048,241
5	39,000,342	39,014,650	39,051,121

Table C.6 The Operating cost of process with 2 washing tank

No. mixer	No. reactor		
	1	2	3
1	39,064,509	39,039,879	39,041,531
2	38,999,133	39,041,525	39,043,595
3	38,996,192	39,042,844	39,045,351
4	39,038,476	39,046,015	39,046,616
5	39,040,187	39,046,952	39,084,196

Table C.7 The payback period of process with 1 washing tank at10% interest rate

No. mixer	No. reactor		
	1	2	3
1	6.70	6.87	7.01
2	6.59	6.84	7.06
3	6.64	6.89	7.12
4	6.80	6.91	7.13
5	6.84	6.94	7.22

Table C.8 The payback period of process with 2 washing tank at10% interest rate

No. mixer	No. reactor		
	1	2	3
1	6.70	6.83	6.89
2	6.59	6.88	6.95
3	6.64	6.95	7.02
4	6.80	7.05	7.08
5	6.84	7.08	7.20

Table C.9 The payback period of process with 1 washing tank at15% interest rate

No. mixer	No. reactor		
	1	2	3
1	8.45	9.28	9.58
2	8.43	9.22	9.68
3	8.63	9.34	9.78
4	8.74	9.36	9.80
5	8.81	9.44	9.98

Table C.10 The payback period of process with 2 washing tank at15% interest rate

No. mixer	No. reactor		
	1	2	3
1	8.95	9.18	9.35
2	8.74	9.31	9.46
3	8.81	9.44	9.59
4	9.13	9.65	9.70
5	9.22	9.71	9.95

Table C.11 The payback period of process with 1 washing tank at18.5% interest rate

No. mixer	No. reactor		
	1	2	3
1	10.94	12.78	13.44
2	10.92	12.64	13.78
3	11.25	12.88	14.08
4	11.50	12.98	14.12
5	11.65	13.10	14.78

Table C.12 The payback period of process with 2 washing tank at18.5% interest rate

No. mixer	No. reactor		
	1	2	3
1	11.99	12.56	12.88
2	11.48	12.85	13.19
3	11.66	13.14	13.46
4	12.42	13.66	13.72
5	12.61	13.86	14.62

C.2 Process with glycerol purification section, No impurity

Table C.13 The investment of process with 1 washing tank

No. mixer	No. reactor		
	1	2	3
1	14,229,760	14,812,705	15,130,643
2	14,337,820	14,888,215	15,206,153
3	14,542,035	14,961,486	15,279,424
4	14,635,035	15,030,594	15,290,011
5	14,687,265	15,086,094	15,435,920

Table C.14 The investment of process with 2 washing tank

No. mixer	No. reactor		
	1	2	3
1	14,523,616	14,866,474	14,961,350
2	14,626,462	14,949,824	15,054,933
3	14,703,565	15,036,547	15,143,896
4	14,822,533	15,177,285	15,220,339
5	14,882,050	15,224,746	15,419,502

Table C.15 The batch time of process with 1 washing tank

No. mixer	No. reactor		
	1	2	3
1	11.75	11.17	11.08
2	12.00	11.17	11.58
3	11.92	11.08	11.92
4	11.92	11.17	11.92
5	11.92	11.25	11.00

Table C.16 The batch time of process with 2 washing tank

No. mixer	No. reactor		
	1	2	3
1	12.25	11.92	12.33
2	12.42	11.92	12.00
3	12.42	11.92	11.83
4	11.92	13.00	12.50
5	12.50	12.92	12.83

Table C.17 The Operating cost of process with 1 washing tank

No. mixer	No. reactor		
	1	2	3
1	40,183,512	40,220,639	40,246,892
2	40,224,985	40,681,252	40,285,290
3	40,380,179	40,244,460	40,286,736
4	40,238,015	40,245,824	40,286,945
5	40,239,046	40,217,354	40,728,065

Table C.18 The Operating cost of process with 2 washing tank

No. mixer	No. reactor		
	1	2	3
1	40,451,951	40,242,584	40,442,235
2	40,237,838	40,244,229	40,282,299
3	40,198,897	40,245,549	39,982,630
4	40,385,180	40,248,720	40,285,321
5	40,386,891	40,285,656	40,246,465

Table C.19 The payback period of process with 1 washing tank at10% interest rate

No. mixer	No. reactor		
	1	2	3
1	2.19	2.39	2.45
2	2.31	2.51	2.48
3	2.39	2.43	2.49
4	2.36	2.44	2.49
5	2.38	2.44	2.62

Table C.20 The payback period of process with 2 washing tank at10% interest rate

No. mixer	No. reactor		
	1	2	3
1	2.44	2.41	2.48
2	2.36	2.42	2.44
3	2.37	2.43	2.38
4	2.43	2.46	2.48
5	2.44	2.48	2.53

Table C.21 The payback period of process with 1 washing tank at15% interest rate

No. mixer	No. reactor		
	1	2	3
1	2.38	2.61	2.38
2	2.52	2.75	2.52
3	2.59	2.64	2.59
4	2.57	2.65	2.57
5	2.58	2.66	2.58

Table C.22 The payback period of process with 2 washing tank at15% interest rate

No. mixer	No. reactor		
	1	2	3
1	2.68	2.62	2.67
2	2.57	2.64	2.67
3	2.58	2.65	2.61
4	2.65	2.68	2.71
5	2.67	2.71	2.78

Table C.23 The payback period of process with 1 washing tank at18.5% interest rate

No. mixer	No. reactor		
	1	2	3
1	2.58	2.78	2.86
2	2.68	2.95	2.89
3	2.78	2.83	2.91
4	2.75	2.84	2.91
5	2.76	2.84	3.10

Table C.24 The payback period of process with 2 washing tank at18.5% interest rate

No. mixer	No. reactor		
	1	2	3
1	2.82	2.80	2.89
2	2.75	2.82	2.86
3	2.76	2.84	2.78
4	2.84	2.88	2.90
5	2.85	2.90	2.98

C.3 Sensitivity of palm oil price of process with purification section, 2.5% Beta carotene and 2.5% FFA at 10% interest rate

Table C.25 The payback period of process with 1 reactor and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.23	2.29	2.65	2.75	3.05	3.45	3.96
2	2.42	2.66	2.95	3.31	3.78	4.40	5.31
3	2.48	2.70	3.00	3.38	3.84	4.49	5.41
4	2.48	2.70	3.00	3.38	3.83	4.49	5.40
5	2.50	2.75	3.04	3.41	3.9	4.56	5.50

Table C.26 The payback period of process with 2 reactors and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.51	2.76	3.06	3.42	3.91	4.58	5.51
2	2.54	2.78	3.08	3.48	3.95	4.61	5.58
3	2.51	2.75	3.05	3.42	3.90	4.56	5.50
4	2.56	2.80	3.11	3.50	4.00	4.66	5.64
5	2.58	2.82	3.13	3.51	4.01	4.70	5.69

Table C.27 The payback period of process with 3 reactors and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.59	2.82	3.14	3.52	4.03	4.71	5.70
2	2.60	2.86	3.18	3.56	4.08	4.78	5.79
3	2.52	2.78	3.06	3.42	3.88	4.50	5.38
4	2.58	2.82	3.13	3.52	4.02	4.71	5.70
5	2.65	2.90	3.22	3.62	4.14	4.85	5.90

Table C.28 The payback period of process with 1 reactor and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.50	2.74	3.04	3.41	3.90	4.55	5.50
2	2.49	2.72	3.02	3.39	3.88	4.51	5.45
3	2.49	2.73	3.02	3.40	3.88	4.51	5.44
4	2.52	2.78	3.08	3.44	3.94	4.56	5.55
5	2.51	2.76	3.08	3.42	3.91	4.56	5.50

Table C.29 The payback period of process with 2 reactors and 2 washing tanks

	% deviation of crude palm oil price						
No. mixer	-9%	-6%	-3%	0%	3%	6%	9%
1	2.52	2.78	3.09	3.46	3.95	4.61	5.59
2	2.55	2.80	3.10	3.48	3.98	4.65	5.62
3	2.56	2.81	3.12	3.50	4.00	4.69	5.68
4	2.60	2.86	3.18	3.58	4.10	4.80	5.82
5	2.60	2.85	3.18	3.58	4.08	4.78	5.80

Table C.30 The payback period of process with 3 reactors and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.56	2.81	3.12	3.50	4.02	4.70	5.69
2	2.55	2.80	3.10	3.49	3.99	4.66	5.62
3	2.60	2.85	3.15	3.55	4.08	4.78	5.78
4	2.60	2.88	3.18	3.58	4.10	4.80	5.82
5	2.66	2.92	3.25	3.65	4.18	4.90	5.95

C.4 Sensitivity of palm oil price of process with purification section, 2.5% Beta carotene and 7.5% FFA at 10% interest rate

Table C.31 The payback period of process with 1 reactor and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.16	2.36	2.60	2.88	3.24	3.71	4.35
2	2.36	2.59	2.86	3.20	3.62	4.20	5.00
3	2.39	2.62	2.90	3.22	3.68	4.25	5.06
4	2.40	2.62	2.90	3.24	3.68	4.27	5.08
5	2.41	2.64	2.92	3.26	3.70	4.29	5.11

Table C.32 The payback period of process with 2 reactors and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.44	2.68	2.96	3.30	3.75	4.34	5.18
2	2.46	2.70	2.99	3.34	3.79	4.39	5.24
3	2.48	2.71	3.00	3.36	3.80	4.41	5.28
4	2.49	2.72	3.05	3.38	3.82	4.42	5.30
5	2.50	2.74	3.04	3.39	3.85	4.46	5.34

Table C.33 The payback period of process with 3 reactors and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.51	2.75	3.04	3.40	3.86	4.48	5.35
2	2.52	2.78	3.08	3.42	3.90	4.52	5.42
3	2.52	2.78	3.08	3.42	3.90	4.53	5.42
4	2.55	2.80	3.10	3.46	3.94	4.58	5.48
5	2.58	2.82	3.12	3.49	3.98	4.61	5.52

Table C.34 The payback period of process with 1 reactor and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.42	2.68	2.96	3.30	3.78	4.38	5.24
2	2.39	2.61	2.89	3.22	3.66	4.22	5.05
3	2.40	2.61	2.90	3.24	3.66	4.22	5.04
4	2.42	2.65	2.92	3.28	3.72	4.30	5.13
5	2.41	2.64	2.92	3.26	3.70	4.28	5.15

Table C.35 The payback period of process with 2 reactors and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.42	2.68	2.94	3.29	3.73	4.32	5.16
2	2.45	2.69	2.96	3.30	3.76	4.35	5.20
3	2.46	2.70	2.98	3.32	3.78	4.39	5.24
4	2.50	2.75	3.02	3.40	3.88	4.49	5.38
5	2.50	2.74	3.02	3.39	3.88	4.48	5.35

Table C.36 The payback period of process with 3 reactors and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.46	2.70	2.98	3.34	3.79	4.40	5.25
2	2.45	2.69	2.96	3.32	3.78	4.36	5.20
3	2.50	2.72	3.02	3.38	3.85	4.46	5.32
4	2.50	2.75	3.04	3.40	3.88	4.48	5.38
5	2.58	2.80	3.10	3.48	3.95	4.59	5.49

C.5 Sensitivity of palm oil price of process with purification section, 2.5% Beta carotene and 12.5% FFA at 10% interest rate

Table C.37 The payback period of process with1 reactor and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	2.99	3.74	3.95	4.72	5.92	8.02	11.79
2	3.18	4.01	4.24	5.10	6.48	8.91	13.78
3	3.21	4.06	4.28	5.16	6.55	9.01	14.04
4	3.22	4.08	4.30	5.18	6.58	9.08	14.15
5	3.25	4.10	4.32	5.22	6.62	9.15	14.30

Table C.38 The payback period of process with 2 reactors and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	3.28	4.14	4.38	5.28	6.70	9.26	14.00
2	3.31	4.18	4.42	5.32	6.79	9.42	14.86
3	3.32	4.20	4.44	5.38	6.82	9.50	15.03
4	3.34	4.22	4.48	5.40	6.88	9.58	15.20
5	3.24	4.06	4.28	5.11	6.40	8.60	12.55

Table C.39 The payback period of process with 3 reactors and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	3.35	4.22	4.48	5.40	6.88	9.59	15.18
2	3.38	4.28	4.52	5.49	7.00	9.80	15.82
3	3.40	4.30	4.55	5.51	7.05	9.88	16.02
4	3.41	4.32	4.57	5.52	7.08	9.91	16.12
5	3.31	4.16	4.38	5.55	6.59	8.92	13.21

Table C.40 The payback period of process with 1 reactor and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	3.22	4.09	4.31	5.20	6.60	9.12	14.24
2	3.22	4.06	4.29	5.16	6.55	9.04	14.05
3	3.22	4.06	4.29	5.16	6.54	8.99	13.84
4	3.28	4.12	4.36	5.28	6.69	9.26	14.52
5	3.28	4.15	4.38	5.29	6.72	9.31	14.62

Table C.41 The payback period of process with 2 reactors and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	3.28	4.14	4.38	5.29	6.71	9.30	14.61
2	3.30	4.18	4.40	5.32	6.78	9.40	14.80
3	3.32	4.20	4.44	5.36	6.82	9.50	15.00
4	3.36	4.25	4.49	5.42	6.91	9.65	15.36
5	3.38	4.28	4.52	5.48	7.00	9.80	15.85

Table C.42 The payback period of process with 3 reactors and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	3.18	3.99	4.50	5.02	6.28	8.84	12.28
2	3.32	4.20	4.45	5.38	6.86	9.58	15.28
3	3.35	4.24	4.48	5.42	6.91	9.66	15.48
4	3.38	4.28	4.50	5.45	6.98	9.75	15.70
5	3.42	4.32	4.58	5.55	7.08	9.88	15.89

C.6 Sensitivity of palm oil price of process with purification section, No impurity at 10% interest rate

Table C.43 The payback period of process with 1 reactor and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	1.79	1.90	2.04	2.19	2.38	2.59	2.85
2	1.87	2.00	2.14	2.30	2.50	2.72	3.00
3	1.90	2.01	2.18	2.34	2.53	2.78	3.06
4	1.91	2.05	2.19	2.35	2.55	2.80	3.08
5	1.92	2.05	2.20	2.38	2.58	2.80	3.09

Table C.44 The payback period of process with 2 reactors and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	1.94	2.08	2.21	2.39	2.58	2.82	3.11
2	1.95	2.08	2.22	2.40	2.60	2.85	3.14
3	1.96	2.09	2.24	2.41	2.62	2.86	3.15
4	1.98	2.10	2.26	2.42	2.62	2.88	3.18
5	1.98	2.10	2.26	2.44	2.64	2.88	3.18

Table C.45 The payback period of process with 3 reactors and 1 washing tank

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	1.99	2.12	2.28	2.45	2.65	2.90	3.20
2	2.00	2.12	2.29	2.46	2.68	2.92	3.22
3	2.01	2.15	2.30	2.48	2.69	2.94	3.25
4	2.01	2.15	2.30	2.49	2.70	2.95	3.25
5	2.04	2.18	2.32	2.51	2.71	2.98	3.28

Table C.46 The payback period of process with 1 reactor and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	1.98	2.11	2.26	2.44	2.66	2.90	3.20
2	1.91	2.05	2.19	2.36	2.55	2.80	3.08
3	1.92	2.05	2.19	2.36	2.56	2.80	3.08
4	1.94	2.08	2.21	2.39	2.59	2.82	3.12
5	1.98	2.10	2.26	2.42	2.64	2.90	3.20

Table C.47 The payback period of process with 2 reactors and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	1.95	2.08	2.22	2.40	2.60	2.84	3.12
2	1.95	2.09	2.24	2.41	2.61	2.86	3.15
3	1.98	2.10	2.26	2.42	2.63	2.88	3.18
4	2.01	2.15	2.30	2.49	2.70	2.96	3.26
5	2.02	2.16	2.31	2.50	2.72	2.98	3.30

Table C.48 The payback period of process with 3 reactors and 2 washing tanks

No. mixer	% deviation of crude palm oil price						
	-9%	-6%	-3%	0%	3%	6%	9%
1	1.98	2.10	2.25	2.42	2.62	2.86	3.18
2	1.98	2.11	2.26	2.44	2.64	2.90	3.20
3	1.99	2.12	2.28	2.46	2.68	2.91	3.20
4	2.02	2.16	2.32	2.50	2.72	2.98	3.30
5	2.05	2.19	2.35	2.54	2.75	3.00	3.32

APPENDIX D

Gantt Chart

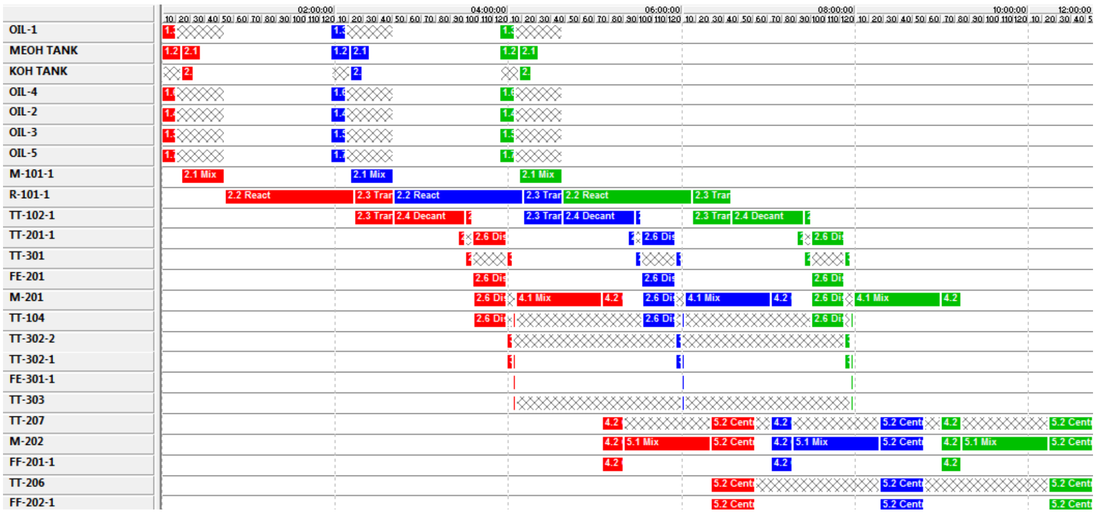


Figure D.1 The Gantt chart of process which had 1 reactor, 1 mixer and 1 washing tank

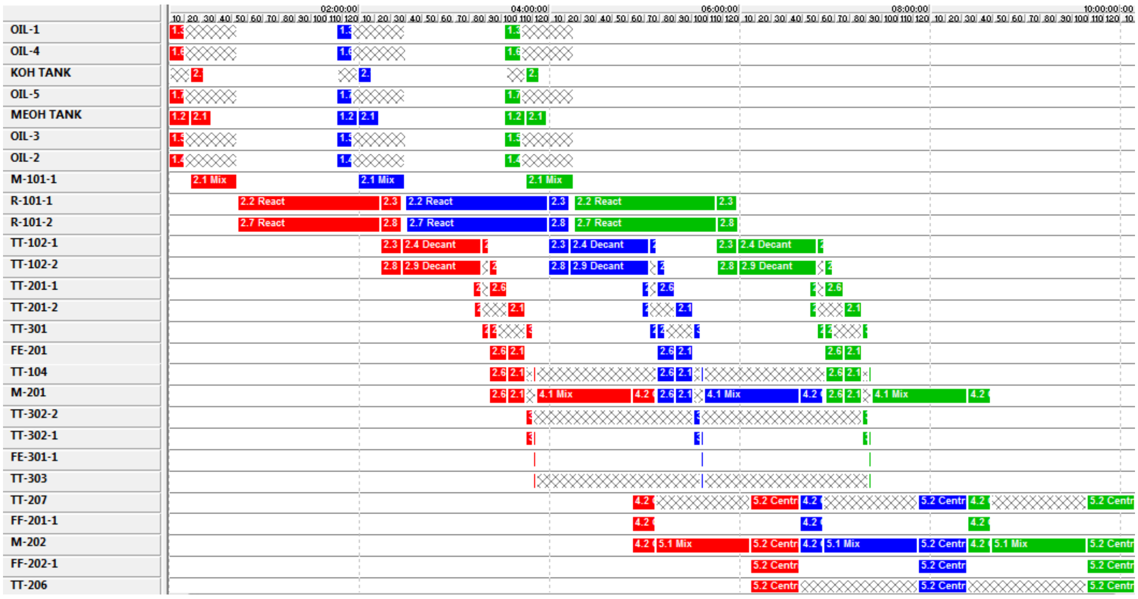


Figure D.2 The Gantt chart of process which had 2 reactors, 1 mixer and 1 washing tank

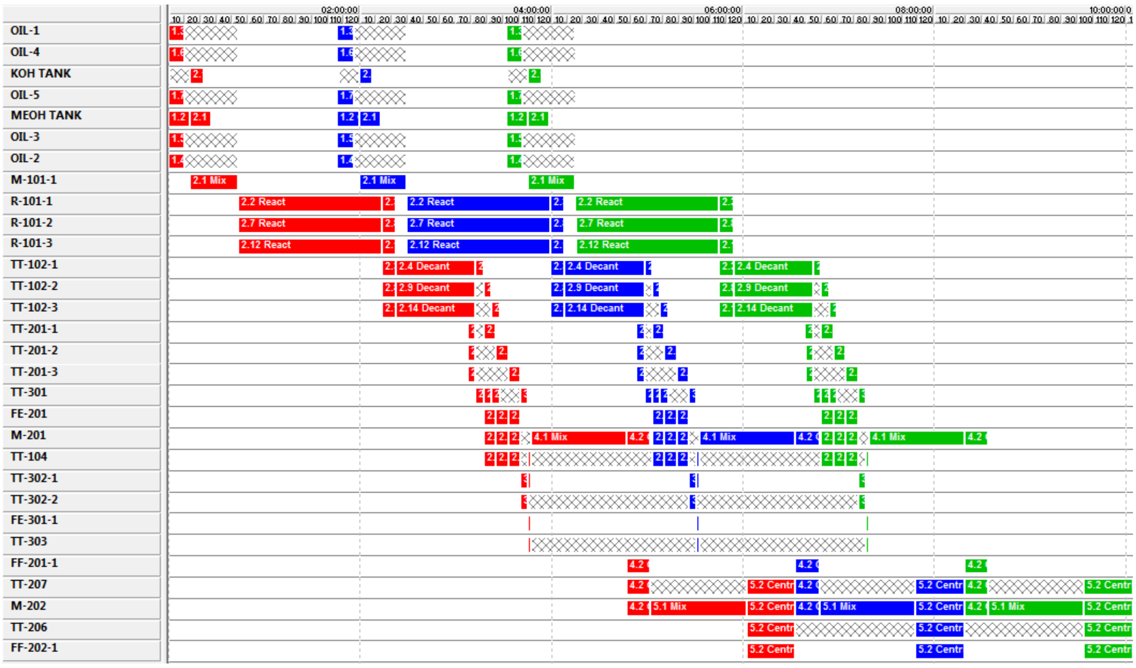


Figure D.3 The Gantt chart of process which had 3 reactors, 1 mixer and 1 washing tank

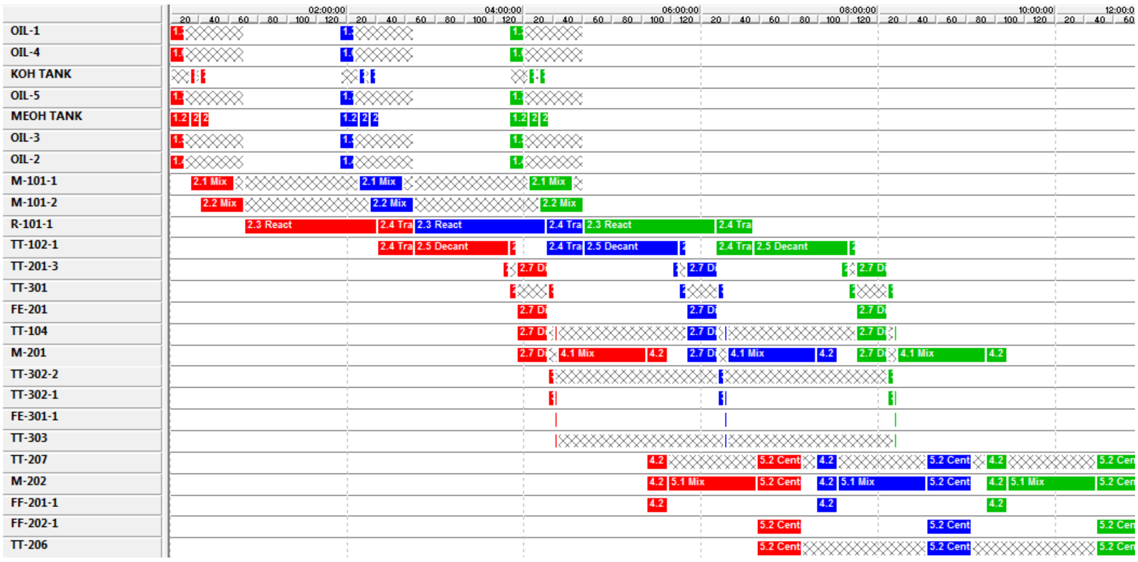


Figure D.4 The Gantt chart of process which had 1 reactor, 2 mixers and 1 washing tank

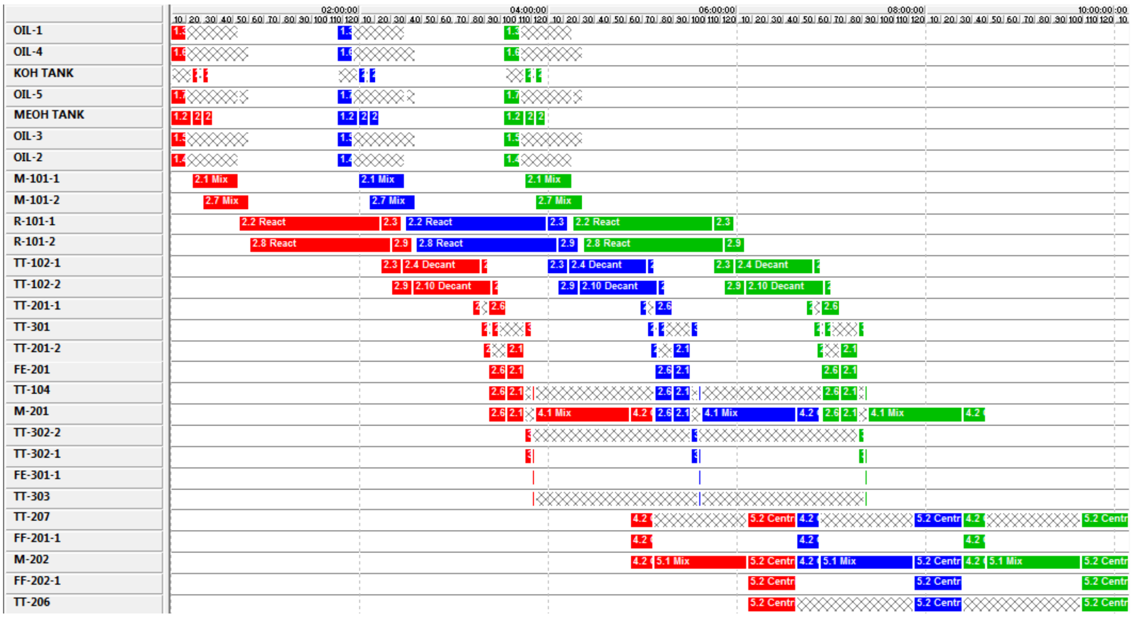


Figure D.5 The Gantt chart of process which had 2 reactors, 2 mixers and 1 washing tank

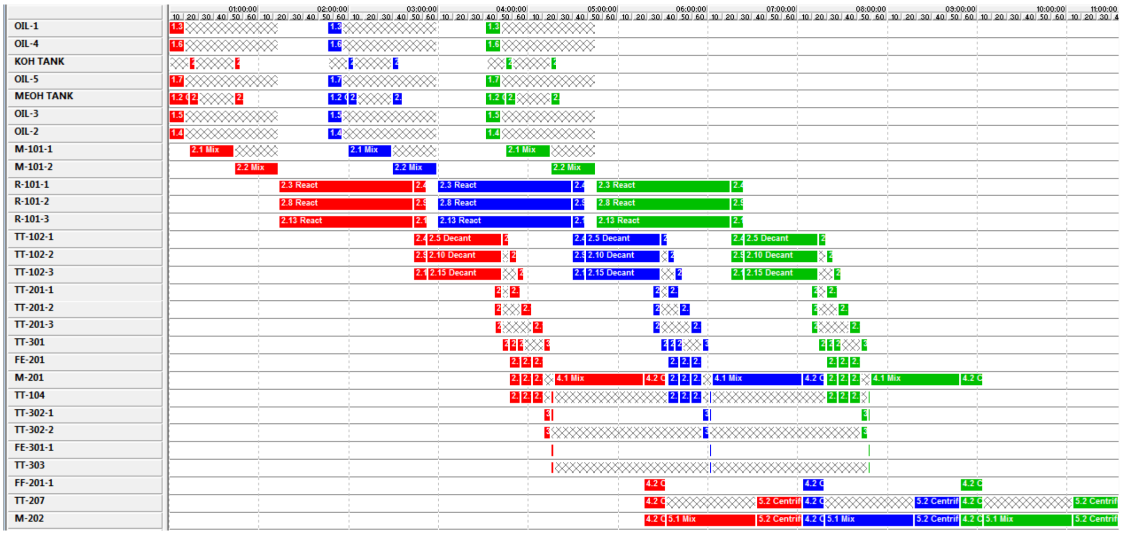


Figure D.6 The Gantt chart of process which had 3 reactors, 2 mixers and 1 washing tank

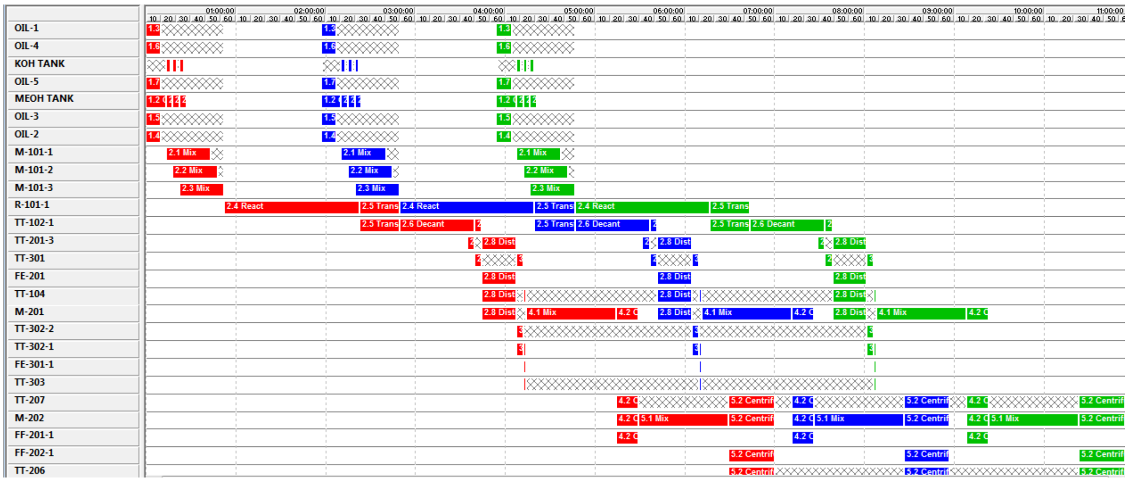


Figure D.7 The Gantt chart of process which had 1 reactor, 3 mixers and 1 washing tank

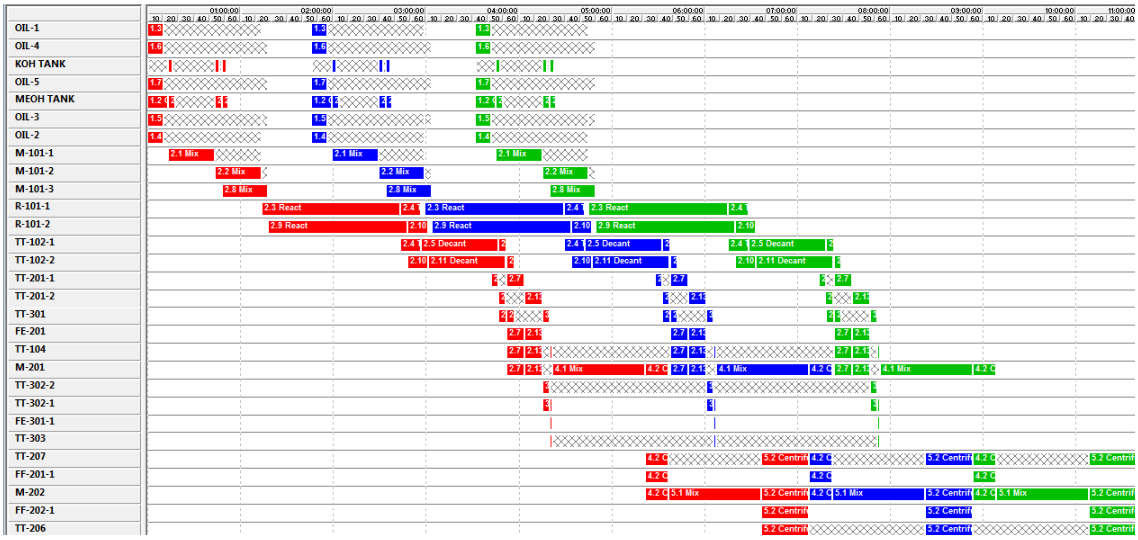


Figure D.8 The Gantt chart of process which had 2 reactors, 3 mixers and 1 washing tank

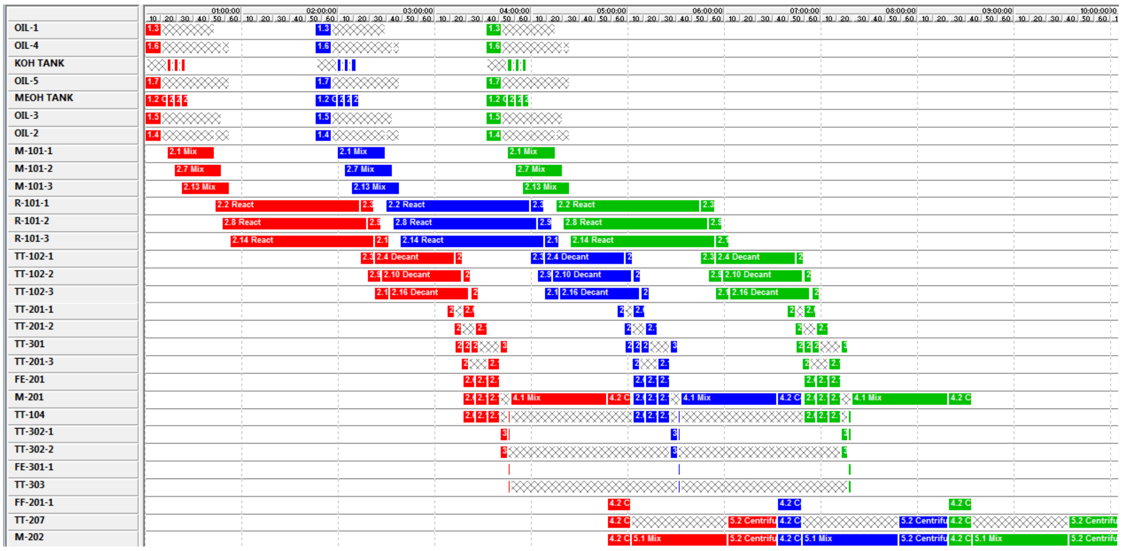


Figure D.9 The Gantt chart of process which had 3 reactors, 3 mixers and 1 washing tank

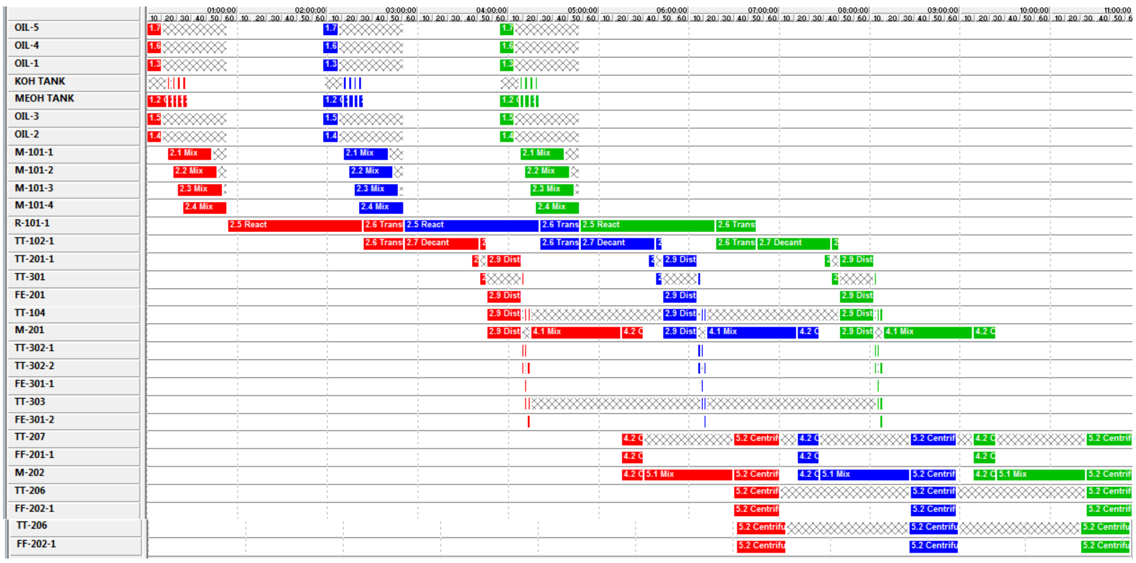


Figure D.10 The Gantt chart of process which had 1 reactor, 4 mixers and 1 washing tank

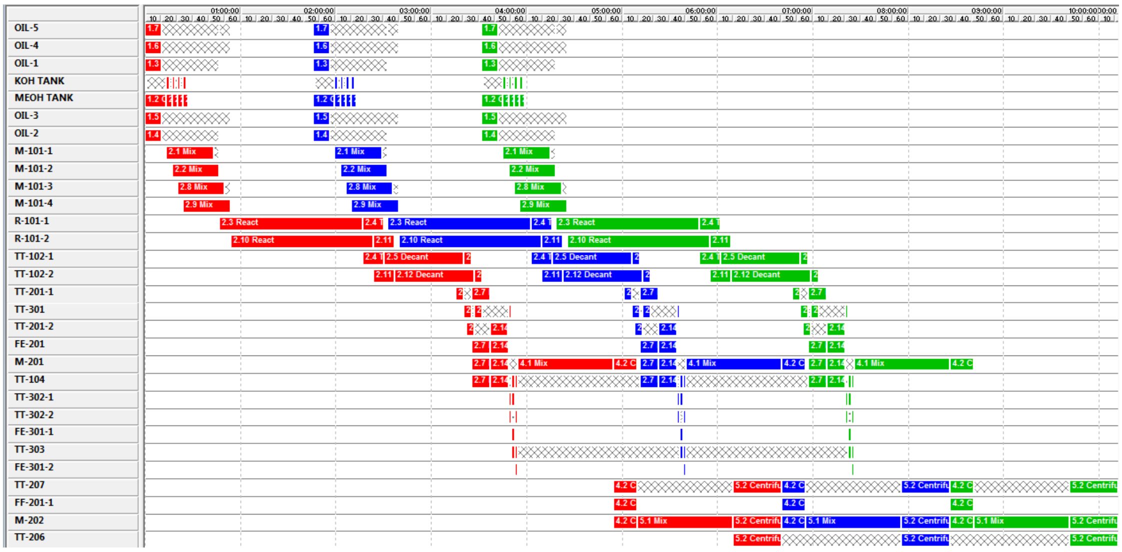


Figure D.11 The Gantt chart of process which had 2 reactors, 4 mixers and 1 washing tank

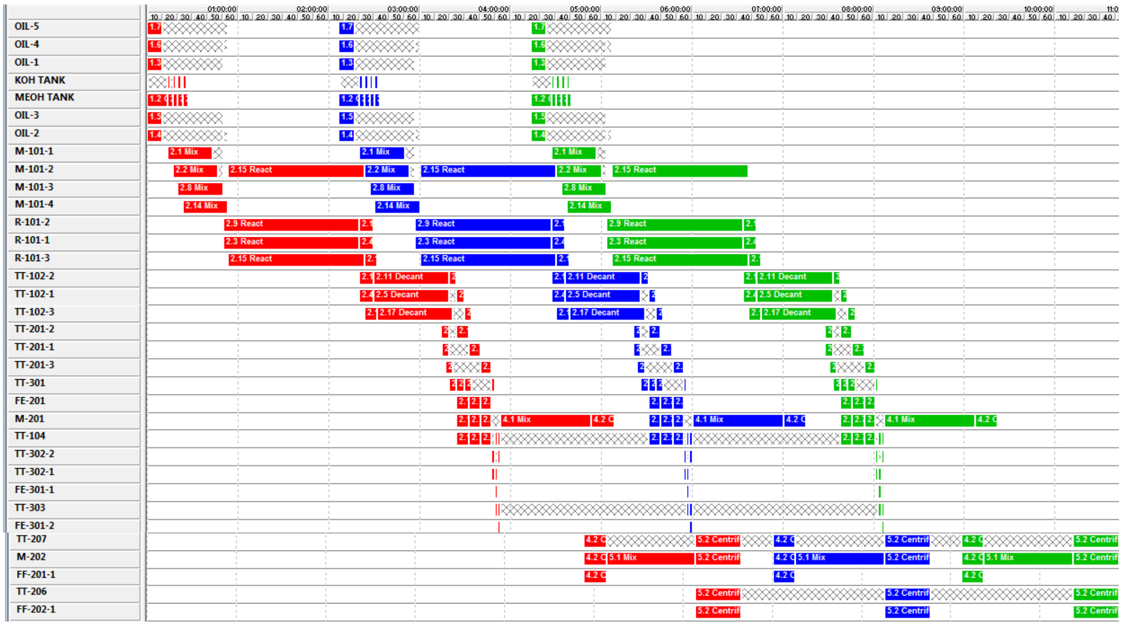


Figure D.12 The Gantt chart of process which had 3 reactors, 4 mixers and 1 washing tank

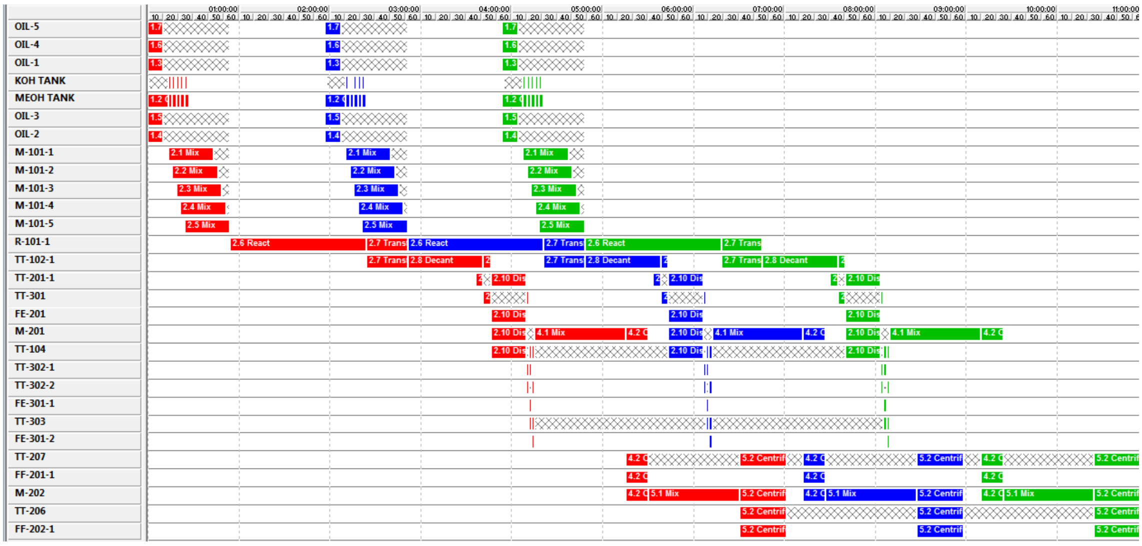


Figure D.13 The Gantt chart of process which had 1 reactor, 5 mixers and 1 washing tank

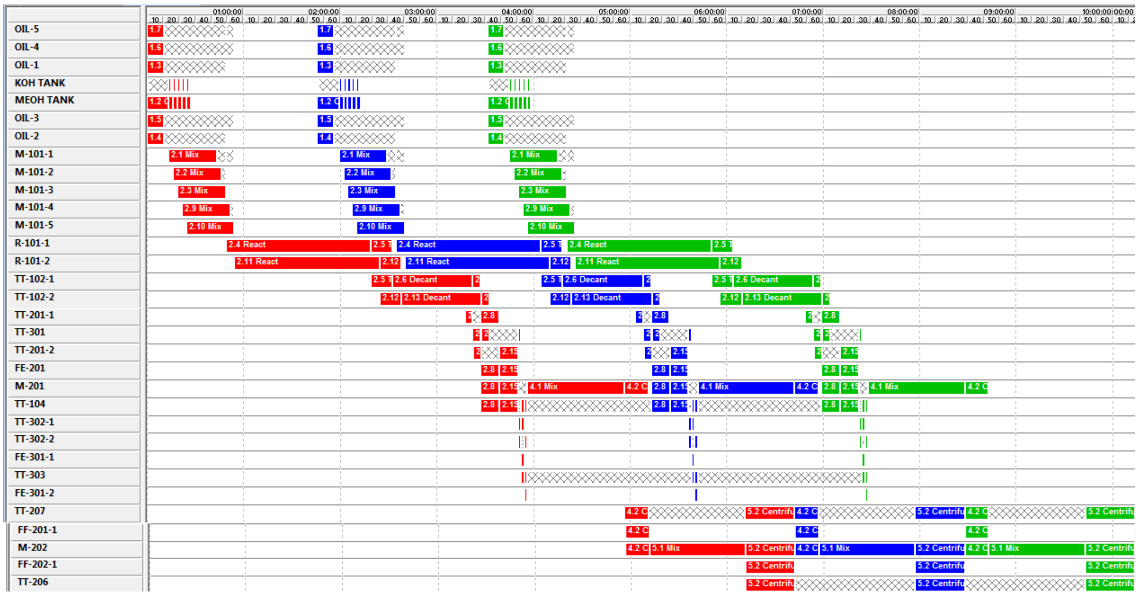


Figure D.14 The Gantt chart of process which had 2 reactors, 5 mixers and 1 washing tank

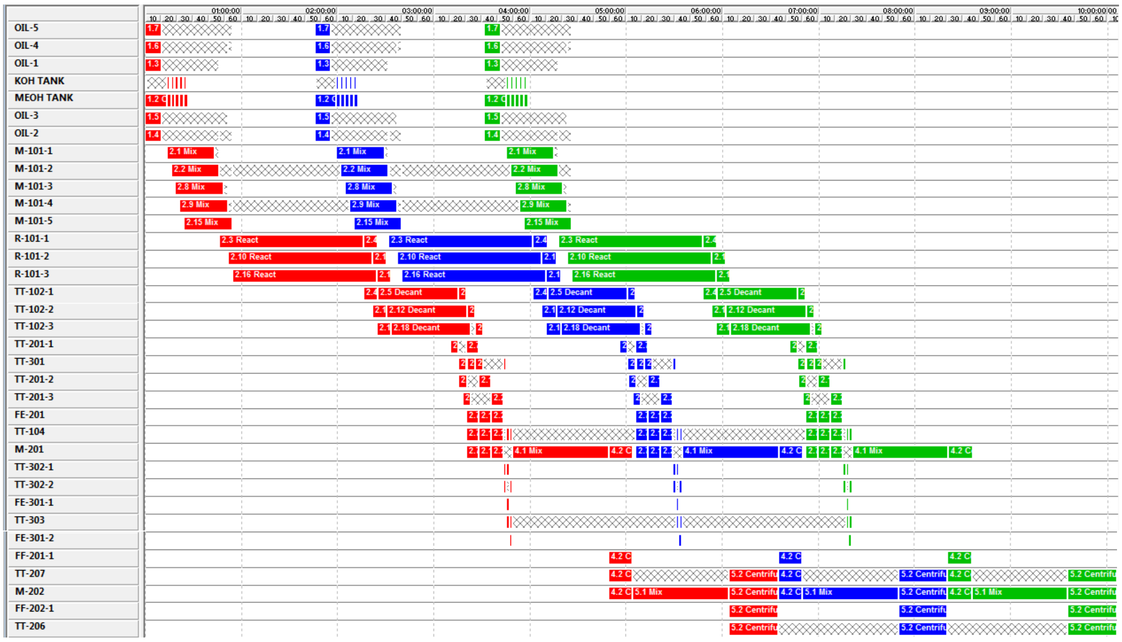


Figure D.15 The Gantt chart of process which had 3 reactors, 5 mixers and 1 washing tank

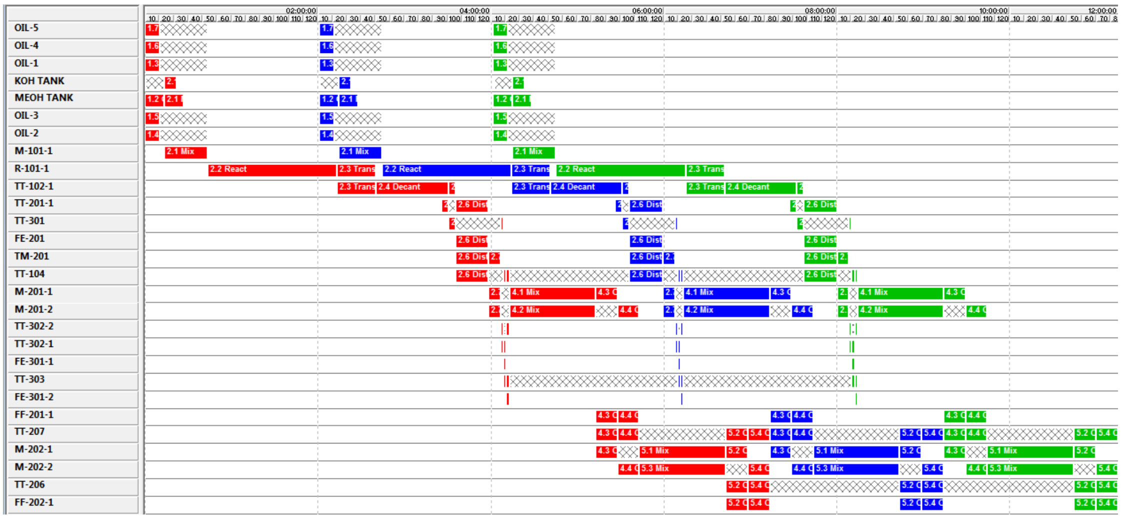


Figure D.16 The Gantt chart of process which had 1 reactor, 1 mixer and 2 washing tanks

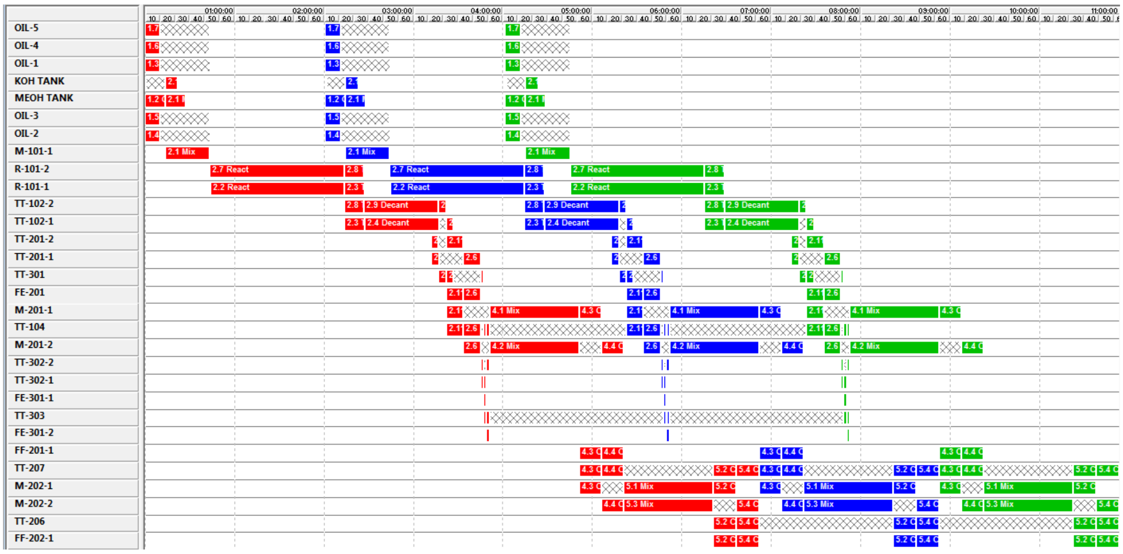


Figure D.17 The Gantt chart of process which had 2 reactors, 1 mixer and 2 washing tanks

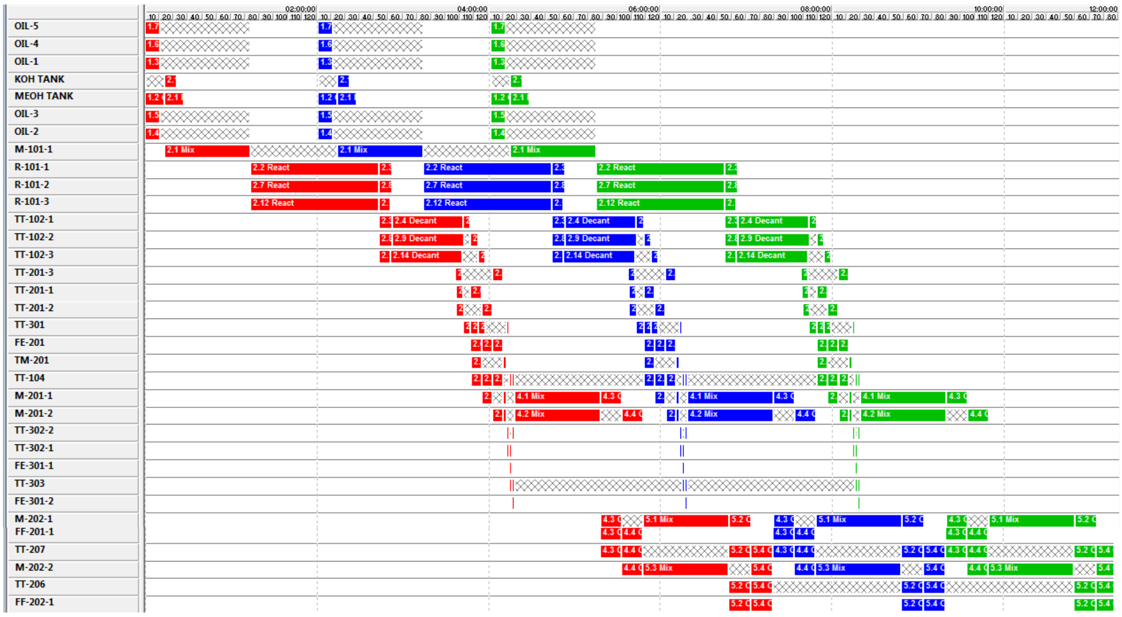


Figure D.18 The Gantt chart of process which had 3 reactors, 1 mixer and 2 washing tanks

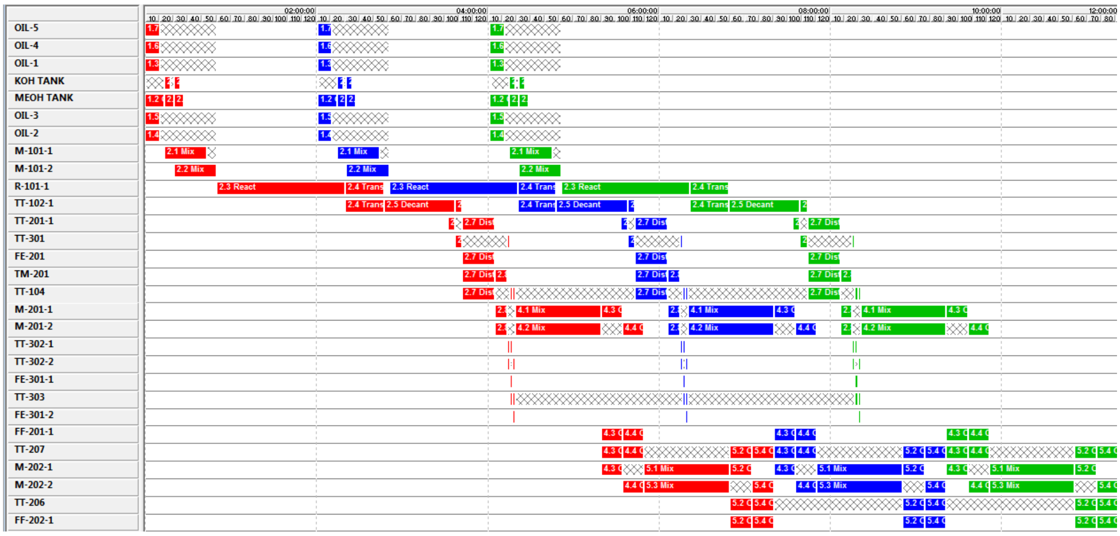


Figure D.19 The Gantt chart of process which had 1 reactor, 2 mixers and 2 washing tanks

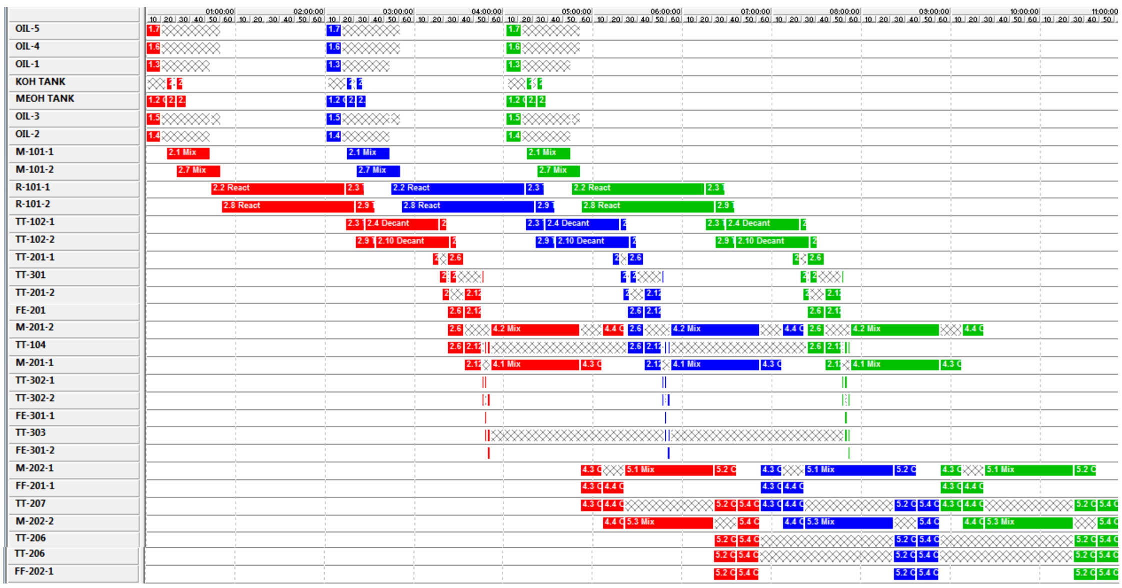


Figure D.20 The Gantt chart of process which had 2 reactors, 2 mixers and 2 washing tanks

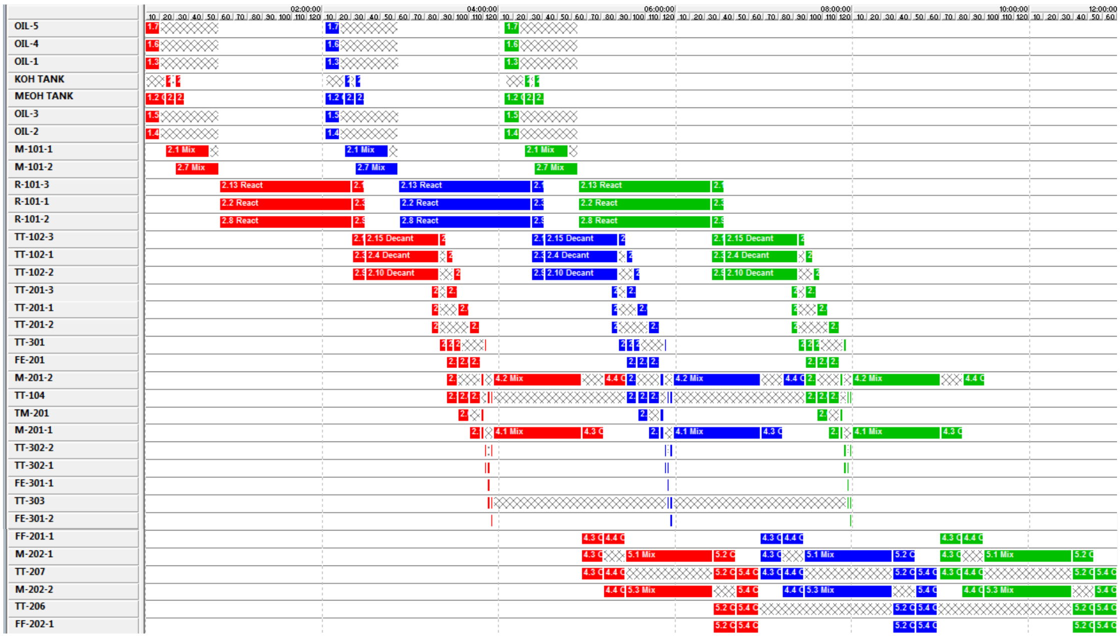


Figure D.21 The Gantt chart of process which had 3 reactors, 2 mixers and 2 washing tanks

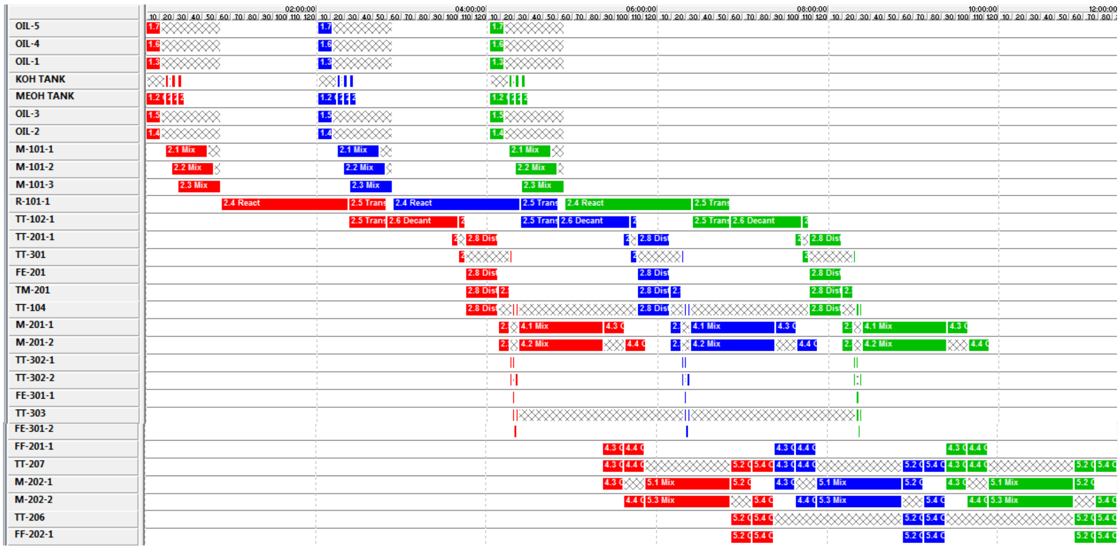


Figure D.22 The Gantt chart of process which had 1 reactor, 3 mixers and 2 washing tanks

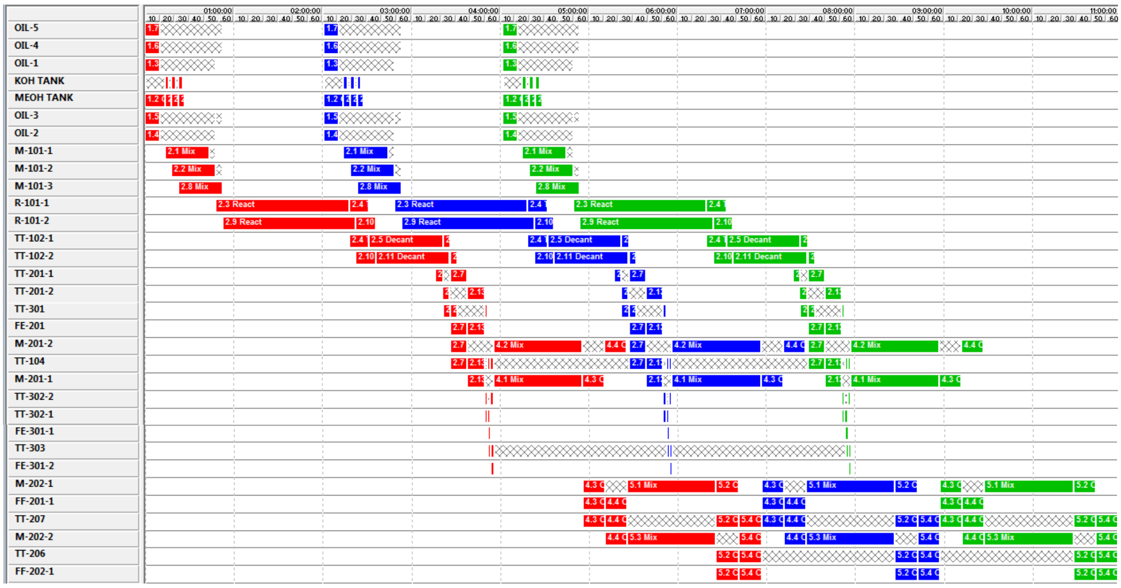


Figure D.23 The Gantt chart of process which had 2 reactors, 3 mixers and 2 washing tanks

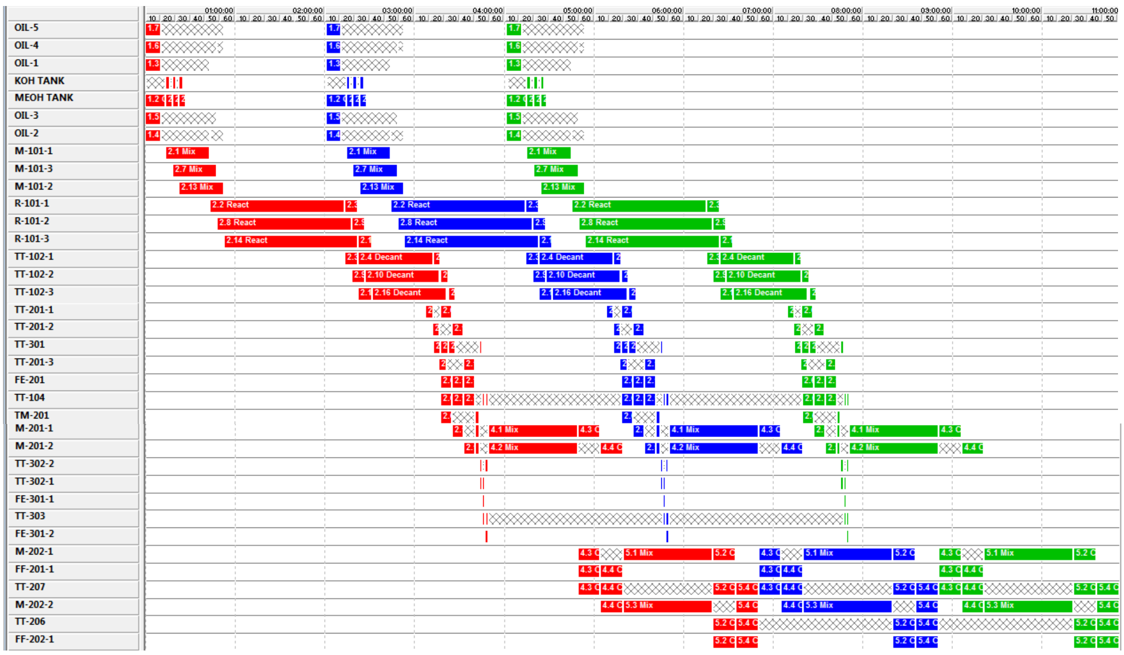


Figure D.24 The Gantt chart of process which had 3 reactors, 3 mixers and 2 washing tanks

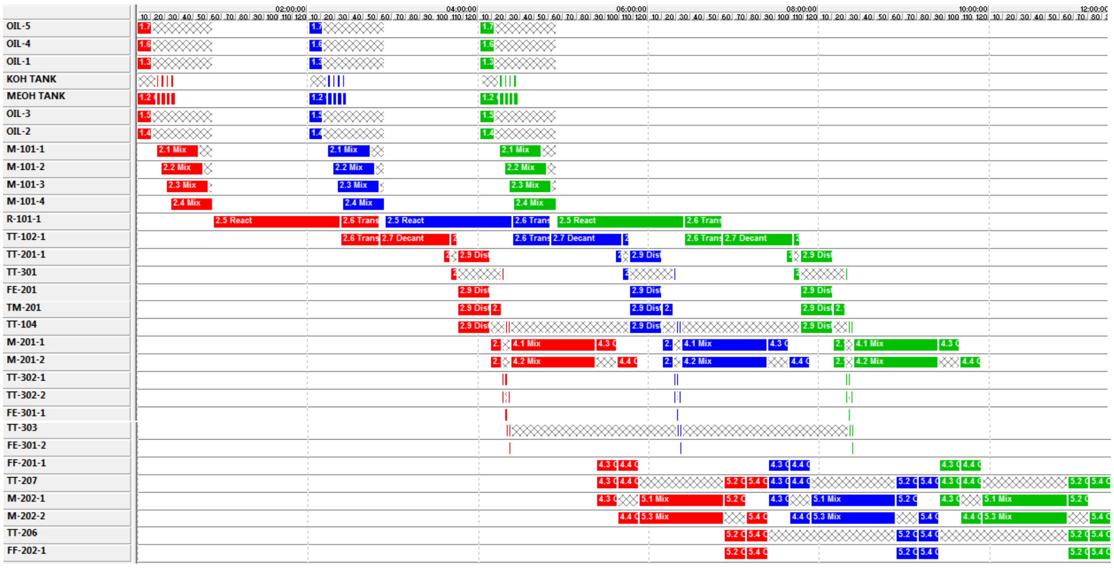


Figure D.25 The Gantt chart of process which had 1 reactor, 4 mixers and 2 washing tanks

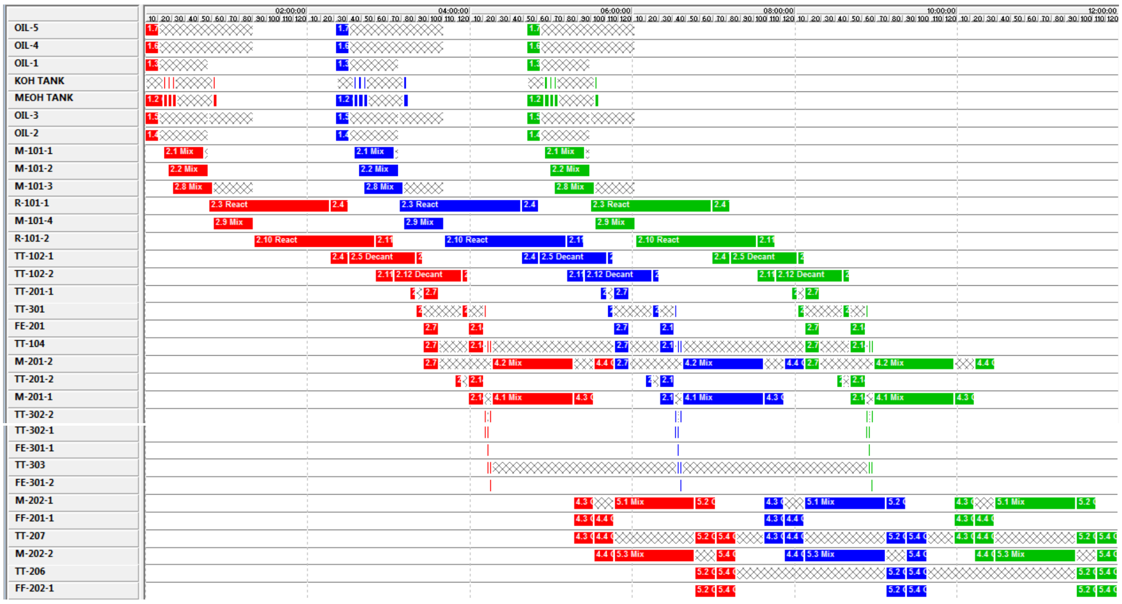


Figure D.26 The Gantt chart of process which had 2 reactors, 4 mixers and 2 washing tanks

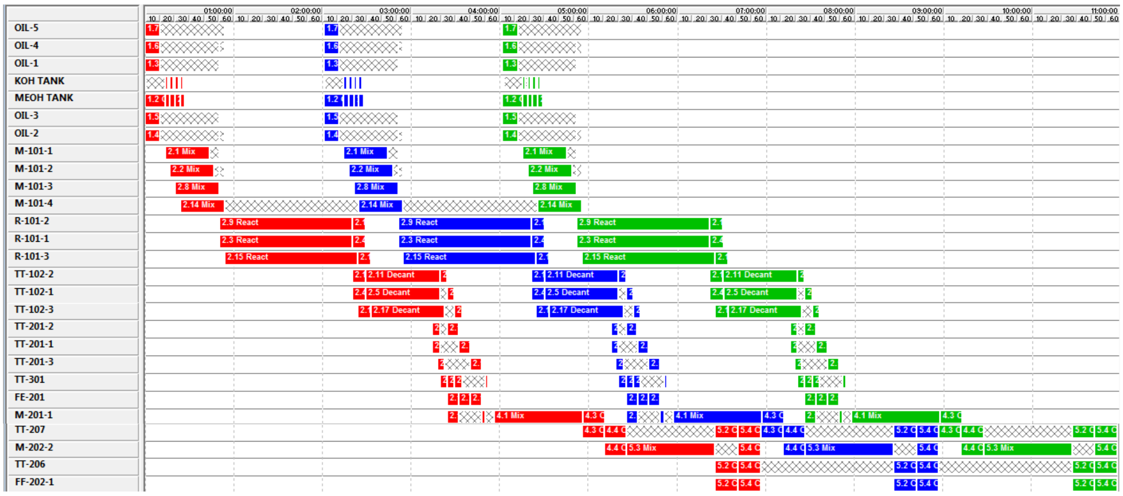


Figure D.27 The Gantt chart of process which had 3 reactors, 4 mixers and 2 washing tanks

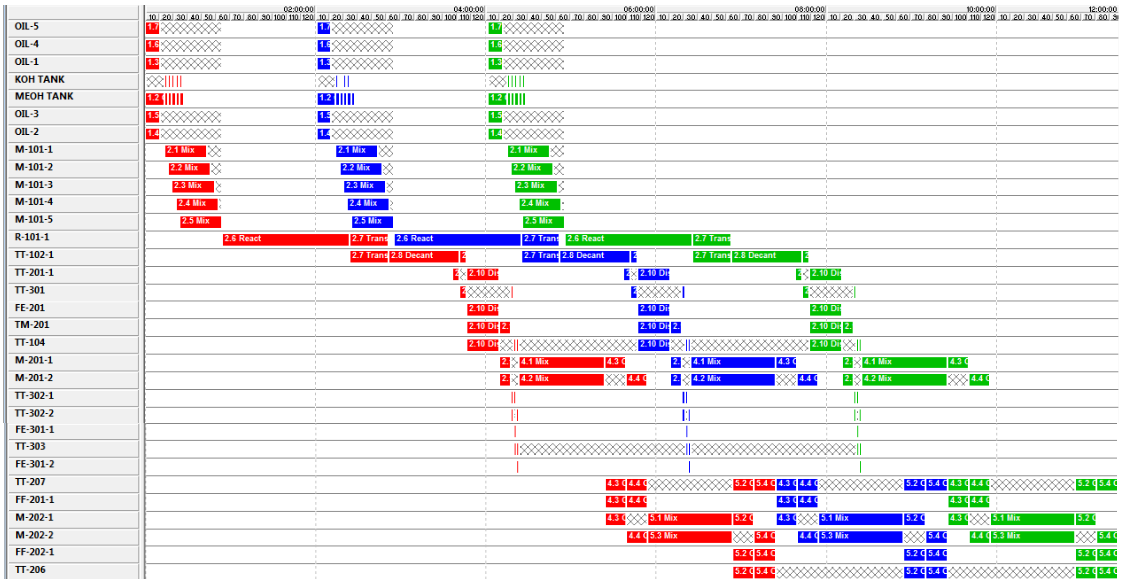


Figure D.28 The Gantt chart of process which had 1 reactor, 5 mixers and 2 washing tanks

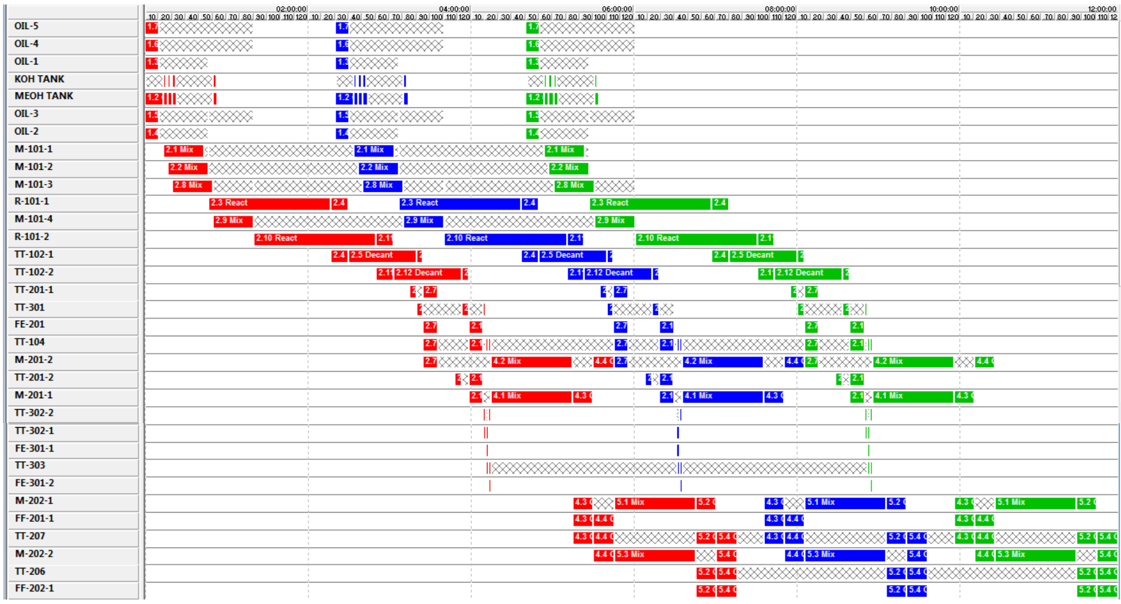


Figure D.29 The Gantt chart of process which had 2 reactors, 5 mixers and 2 washing tanks

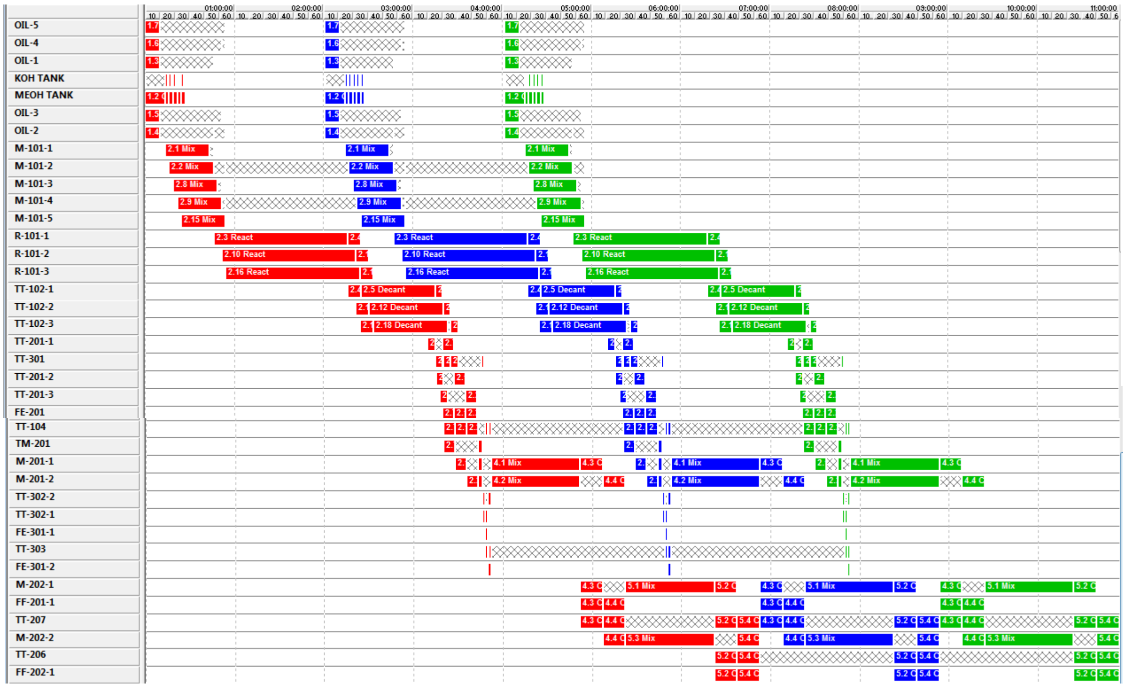


Figure D.30 The Gantt chart of process which had 3 reactors, 5 mixers and 2 washing tanks

CURRICULUM VITAE

NAME Miss Kanyaporn Lertwimolkasem

DATE OF BIRTH 15 September 1987

EDUCATIONAL RECORD

HIGH SCHOOL	High School Graduation Chonburi Sukkhabot School, Chonburi, 2006
BACHELOR’S DEGREE	Bachelor of Engineering (Chemical Engineering) King Mongkut's University of Technology Thonburi, 2009
MASTER’S DEGREE	Master of Engineering (Chemical Engineering) King Mongkut’s University of Technology Thonburi, 2011