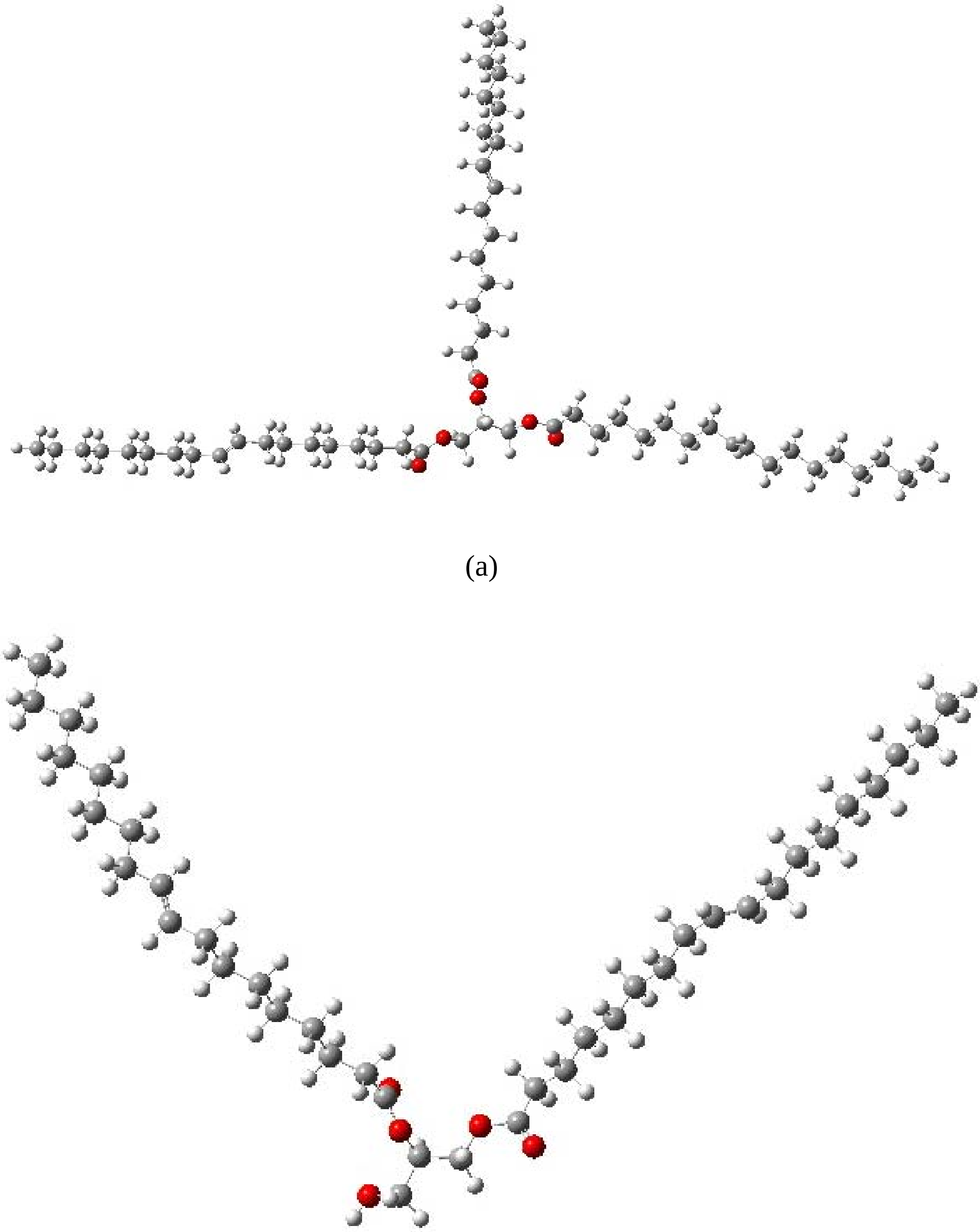
RESULTS AND DISCUSSION

This thesis can be divided into two parts. The first part concerns about improvement of biodiesel production from conventional to reactive distillation in order to propose better performance of biodiesel production. The second part concerns about part studies the dynamic behaviors of the reactive distillation.

1. Structure of components

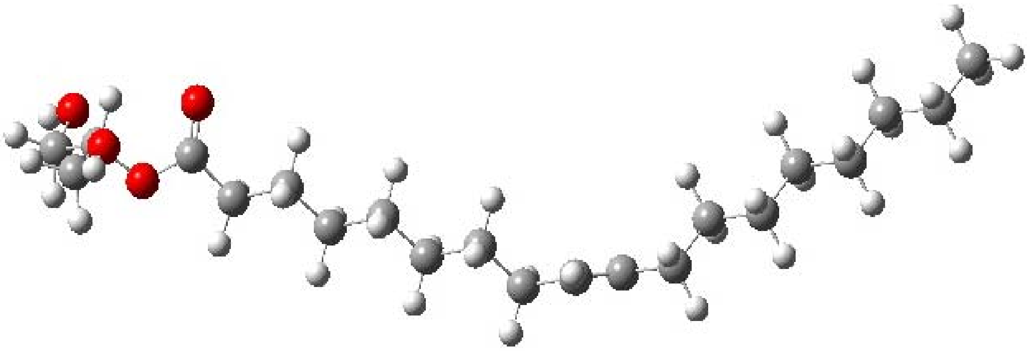
Due to unavailable structure of reactants and products in the library of ASPEN PLUS, these components which were trioleic, dioleic, monooleic, methyl oleate, trilinoleic, dilinoleic, monolinoleic and methyl linoleate were draw by GaussViewW. After drawing, the structures were optimized by GAUSSIAN 03W in order to obtain the stable structure. Figure 14 and 15 demonstrates the structure of mono, di, tri and methyl ester of oleic and linoleic from GAUSSIAN 03W. All of these structures were used in ASPEN PLUS for finding the parameter such as boiling point, molecular weight, heat capacity, coefficients for antoine equation.



(a)

(b)

Figure 14 The structure of (a) trioleic (b) dioleic from GAUSSIAN 03W



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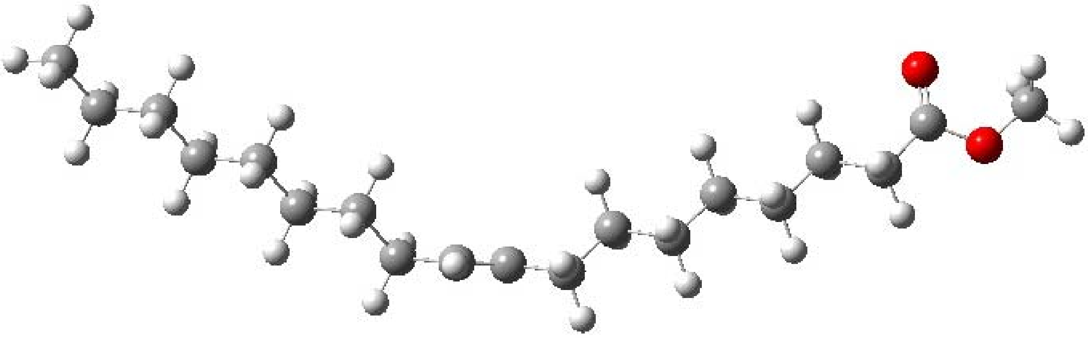
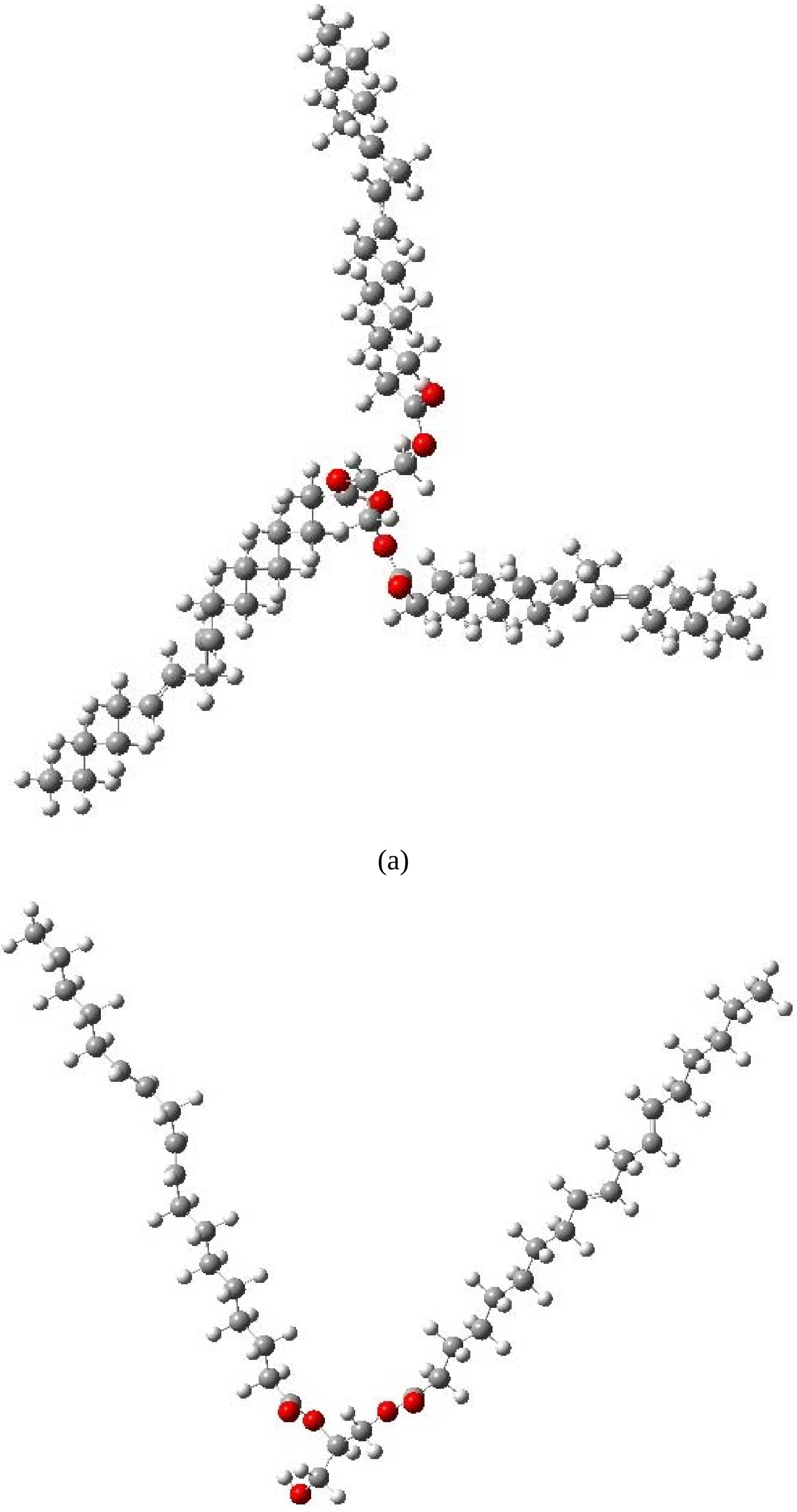


(d)

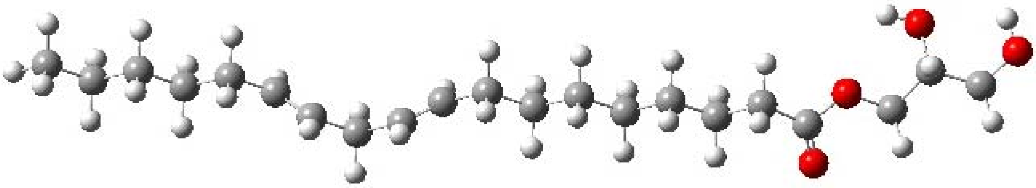
Figure 14 (cont’d) The structure of (c) monoleic (d) methyl oleic from GAUSSIAN 03W



(a)

(b)

Figure 15 The structure of (a) trilinoleic (b) dilinoleic from GAUSSIAN 03W



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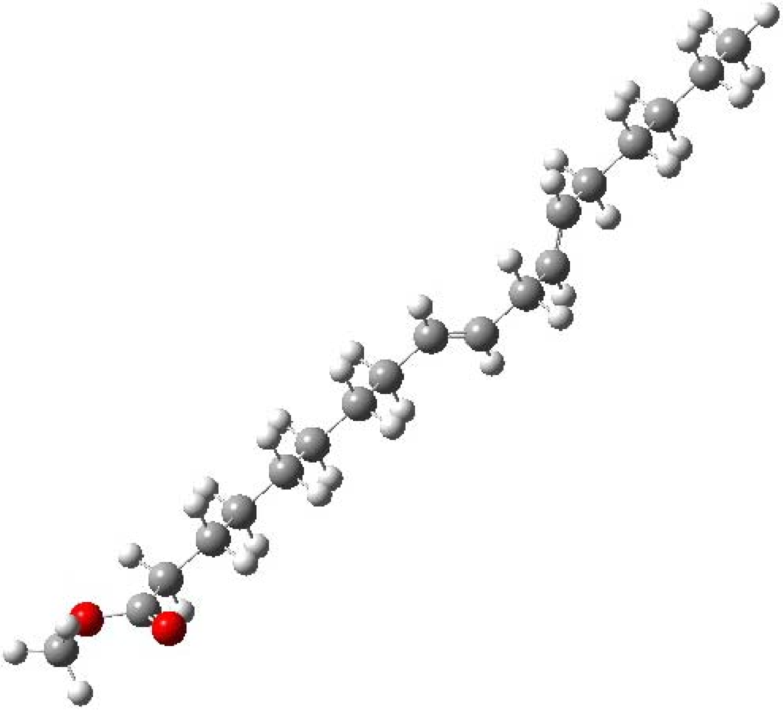


(d)

Figure 15 (cont'd) The structure of (c) monoleic (d) methyl linoleic from GAUSSIAN 03W

2. **The Conventional biodiesel production**

According to Zhang’s work (Zhang et al., 2003), there were four simulations of biodiesel production. Two of them were alkali-catalyzed process, one using virgin oil and other using waste cooking oil. The remaining two were acid-catalyzed processes using waste cooking oil as the raw material. Because this research focuses on the alkali catalyst, the alkali catalyst process of Zhang will be used for applying the reactive distillation. Figure 16 shows the alkali catalyst process to produce biodiesel from virgin oil. In this process consists of 7 processes which are transesterification, methanol recovery, water washing, biodiesel purification, alkali removal, glycerin purification and waste treatment. Normally, after complete reaction, there are two layers between biodiesel and glycerol (Zhou *et al*, 2006). The water washing unit was used to separate the biodiesel and glycerol but it is easy to separate by gravity unit introduced by (Krawcyk, 1996). Therefore, the water washing unit of Zhang (2003) was replaced by using the gravitational unit. Figure 17 displays the modification alkali catalyst process from Zhang (2003). As seen in the Figures 17, there are 4 sections which are transesterification, methanol recovery, glycerol and alkali removing. The sequence of biodiesel production is changed after the methanol recovery. The gravity unit was used to remove glycerol and catalyst. The biodiesel quantity between Zhang’s work (Zhang *et al.*, 2003) and this work are 997 kg/h and 998 kg/h, respectively. In this research, the equipment can be reduced from five to two. In addition, less pump and pipe line between the units can be lowered too. The detail of each process was described below.

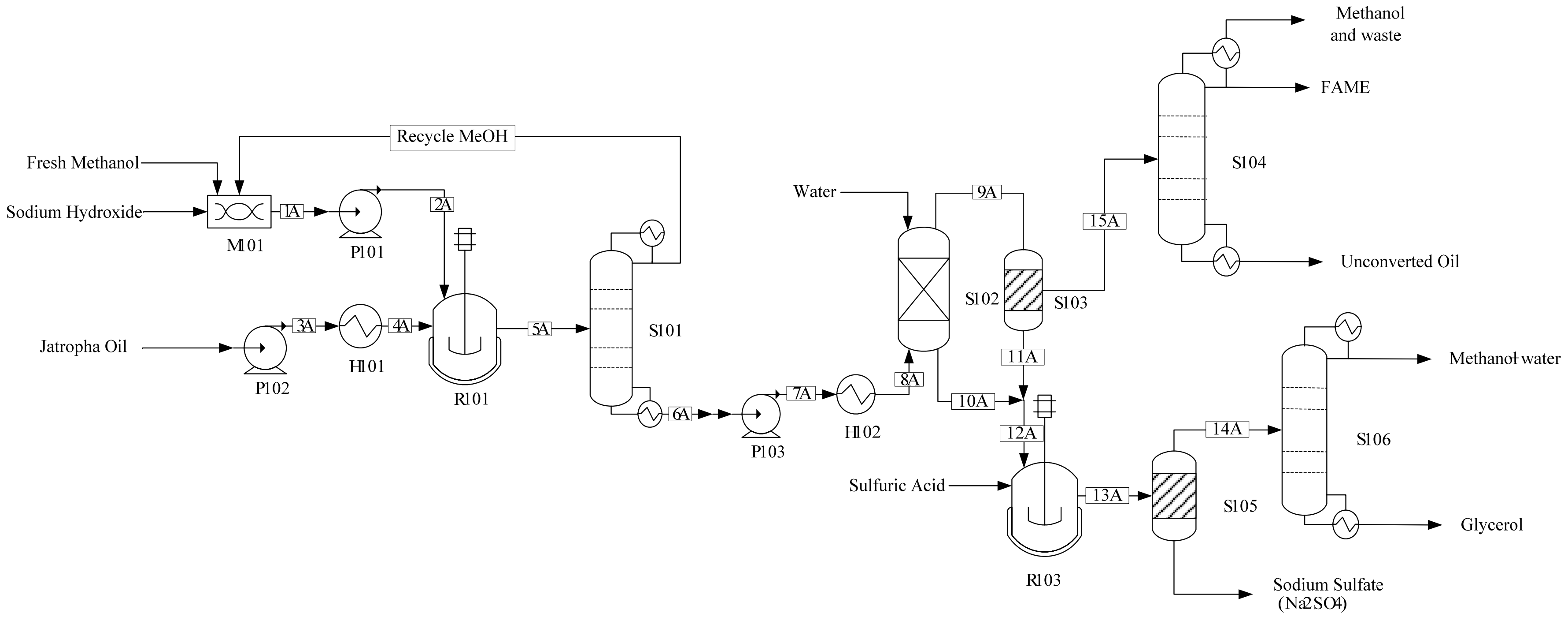


Figure 16 The alkali catalyst process to produce biodiesel from virgin oil

Source: Zhang *et al.* (2003)

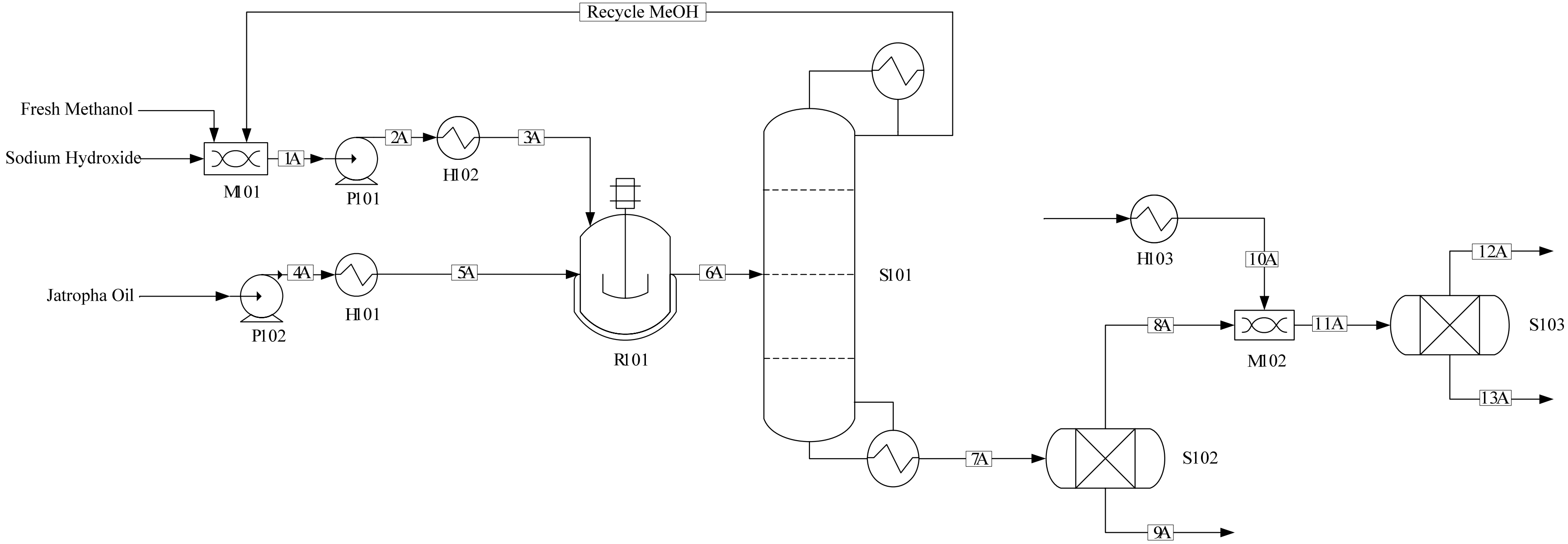


Figure 17 The biodiesel production of this work

2.1 Transesterification

The detail of the transesterification is shown in the Figure 18. The process started with the 1000 kg/hr of Jatropha oil was heated to 60°C before sent into the reactor. At the same time, the sodium hydroxide, 1% wt of Jatropha oil (Freeman *et al*,1984), was mixed with methanol of 1:5 mass ratio or 6:1 molar ratio alcohol to oil (Freeman *et al*,1986) in static mixer M101 to produce 0.5% sodium methoxide which used as a catalyst. The fresh methanol was mixed with recycle methanol before mix with sodium hydroxide. After that, sodium methoxide was sent to the reactor which transesterification was occurred. Due to the detail information on the kinetic is not available, a simple conversion of biodiesel is assumed to be 95% (Zhang *et al*, 2003). But there are a lot of kinetic models of transesterification reaction available (Freedman *et al*., 1986; Noureddini and Zhu, 1997; Darnoko and Cheryan, 2000; Foon *et al*., 2004; Karmee *et al*., 2004; Vicente *et al*., 2005; Vicente *et al*., 2006). Because of the same catalyst selected and similar component of oil, the kinetic model of Noureddini and Zhu (1997) is selected to describe the transesterification reaction of jatropha oil. The excess of methanol is sent to the recovery section in order to recycle the methanol.

The simulation results, the fresh and recycle methanol which were 130 kg/h and 38.4 kg/h respectively was mixed with the sodium hydroxide (12 kg/hr) in the static mixer M101. The mixture called sodium methoxide at rate 180 kg/h was pumped to the heat exchanger in order to increase the temperature to 60°C. The 1,000 kg/h of fresh Jatropha oil which was assumed to trioleic and trilinoleic as a major composition were pumped and preheated with heat exchanger H101 to reach 60 °C before sent to reactor. In the reactor R101, temperature was controlled at 60 ^{o}C, 1 atm pressure and 1 hour operated. The final product is approximately 1,180 kg/h that contain biodiesel (methyl Oleic and methyl linoleic), glycerol, unreacted methanol, catalyst and some of intermediate.

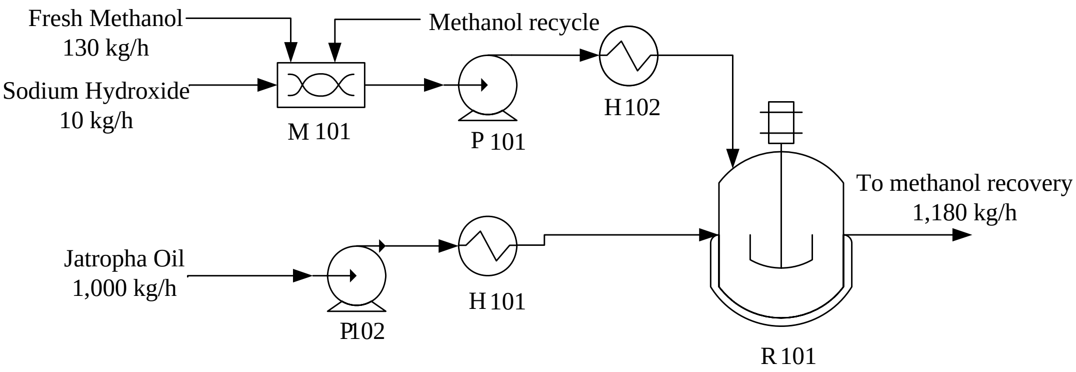


Figure 18 Transesterification section

2.2 Methanol recovery

From transesterification section, the product and unreacted reactant are sent to distillation to remove and recycle unreacted methanol in order to reduce fresh feed of methanol. Figure 19 shows the methanol recovery section which methanol and other components are separated by the distillate and bottom stream respectively. The pressure column should be operated between 0.2 - 0.3 atm. (Zhang *et al.*, 2003) to keep the bottom temperature below 150 ^{o}C because of declining of biodiesel. The designs of distillation are 5 stages, reflux ratio of 1 and boiler ratio of 0.6 to separate methanol from other component.

The 1,180 kg/h from reactor R101 are sent to distillation. The top and bottom pressures of distillation are fixed at 1 atm and the bottom temperature is lower than 150 ^{o}C. Therefore, the operation at the atmospheric pressure is better because higher capital cost come up with high operating pressure (Stitt, 2002). In the recycle stream, 38.4 kg/h of methanol is recycled and mixed with fresh methanol, then charged back into reactor. The 1,142 kg/h of bottom stream is sent to glycerin removing. The quality of methanol in this stream should not higher than 0.2 % (EN 14214). If there is excess methanol, the separation unit should be installed in the down stream.

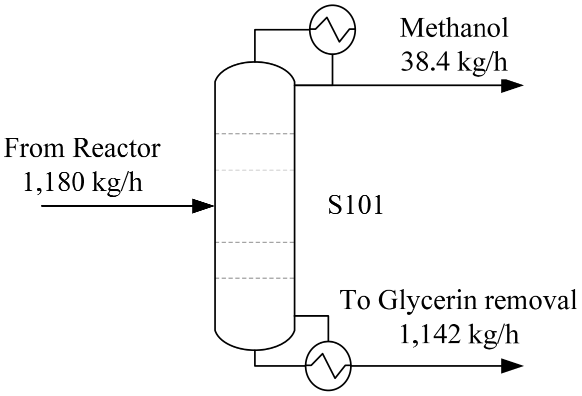


Figure 19 Methanol recovery section

2.3 Glycerin removal

Due to the mass fraction of glycerol in biodiesel should not be over 0.24 % (EN 14214), the removal glycerin unit is significantly. After the methanol recovering section, the water washing column was used to separate the biodiesel from the glycerin (Connemann and Fischer, 1998; Zhang *et al.*, 2003). But it is easy way to remove glycerin from other component instead of using water washing column because glycerin does not dissolve in the biodiesel. There are two layers between biodiesel and glycerin (Zhou *et al*, 2006). The residue curve is applied to study the dissolves of biodiesel and glycerin. Figure 20 shows the residue curve for glycerin, biodiesel and methanol. Glycerin is hardly soluble in biodiesel and the mixtures are two phase. Therefore, the gravity unit is enough to be used for this separation.

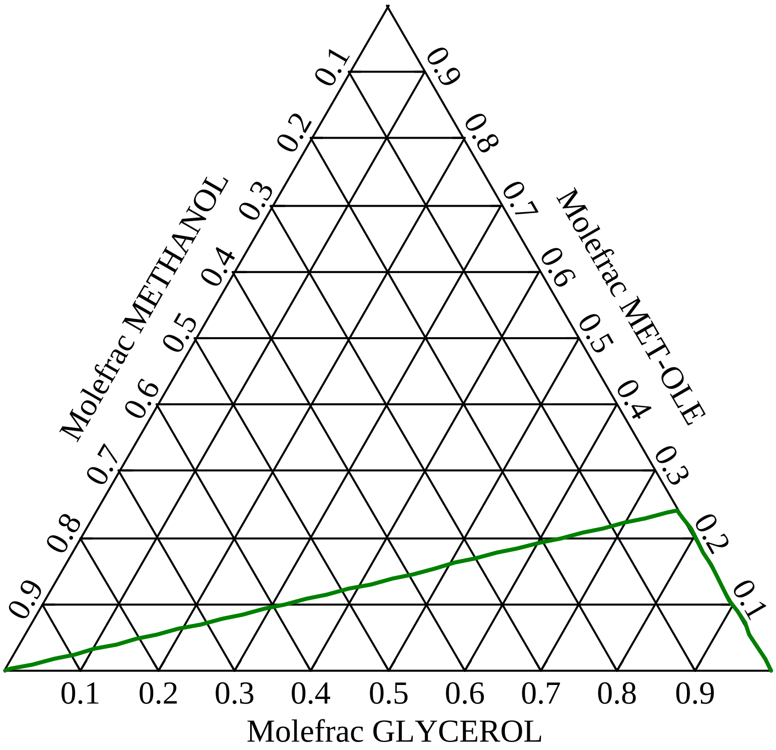


Figure 20 Residue curve for glycerol, biodiesel and methanol

Figure 21 shows the glycerin removal unit (decanter). The bottom stream from distillation is sent to the decanter unit to separate first and second liquid. Glycerin identified as a key component of second liquid was removed in the second liquid stream. The first liquid stream consisting 96.9 % biodiesel was sent to the alkali removal unit. Glycerin containing which is lower than 0.1 % is acceptable with the standard (EN 14214). In the second liquid stream, there are 80.4 % glycerin and only 7 % biodiesel. Therefore, glycerin can be sold as a low grade because of its quality.

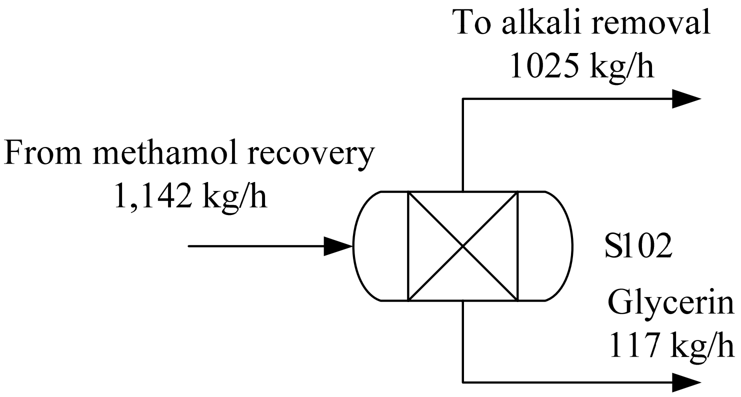


Figure 21 Glycerin removal section

2.4 Alkali removal

The amount of sodium hydroxide should not be higher than 0.8 % (EN 14214). Therefore, the alkali removal section shown in the Figure 22 is used in order to decrease the amount of sodium hydroxide in biodiesel product. First of all, 900 kg/h of hot water (50 ^{o}C) is mixed with the first liquid stream coming from the glycerol removal in the static mixer. Secondly, the mixing component is sent to decant which operates at 50 ^{o}C and 1 atm. The water set as a key component is removed in the second liquid stream. The efficiency of decant is set 100 % because biodiesel and water are not dissolve together. These results can observe in the Figure 20 which shows the residue curve for biodiesel, water and methanol.

As seen in the Figure 23, the biodiesel leaving from glycerin removal is sent to the alkali removal. The 900 kg/h of water which is minimum requirement is used to dissolve sodium hydroxide in the product. In the static mixer, the alkali catalyst is then dissolved into water. The mixtures are sent to decanter used for removing between biodiesel and water. The biodiesel 97.7 % at rate 982 kg/h is obtained. The quality of biodiesel meets the standard (EN 12142).

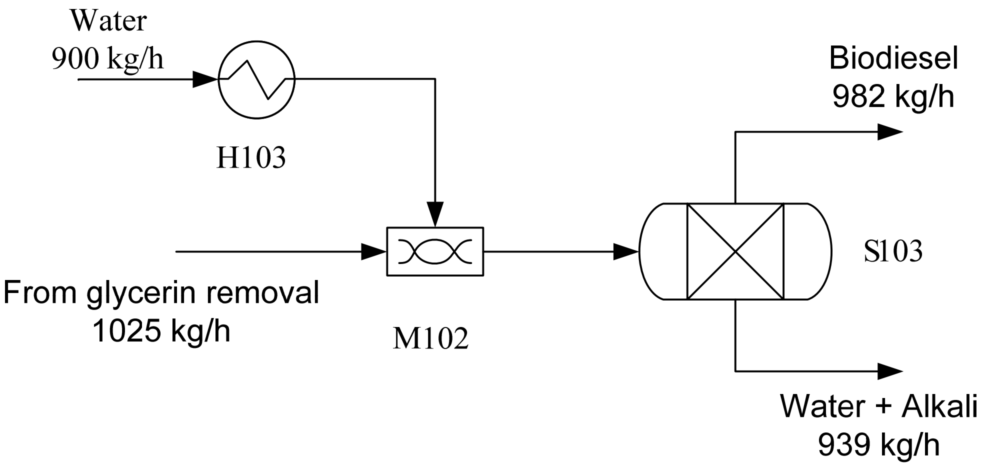


Figure 22 Alkali removal section

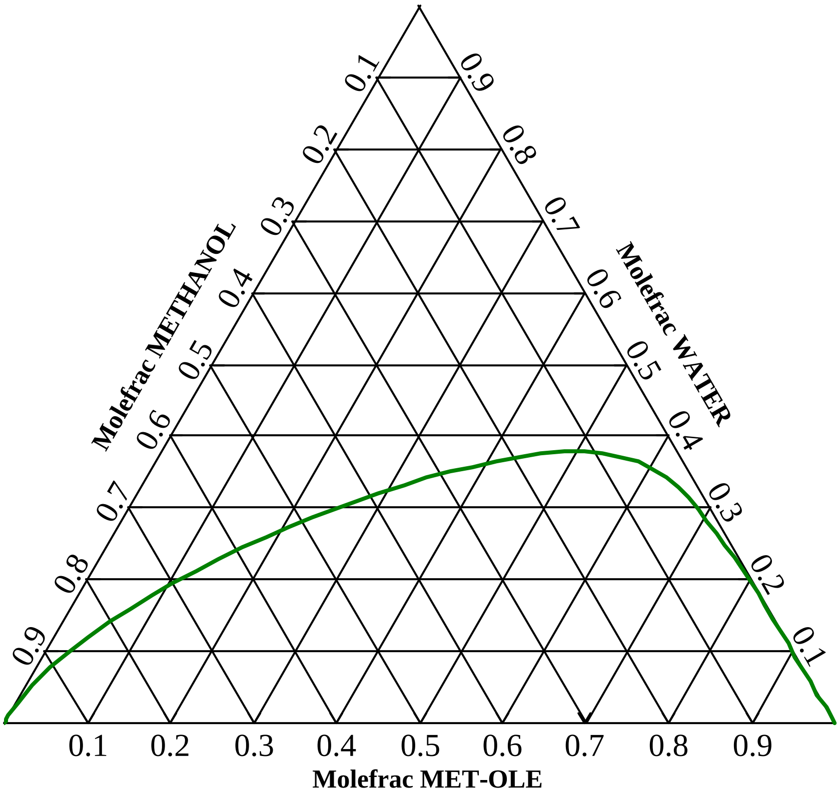


Figure 23 Residue curve for biodiesel, water and methanol

3. The Biodiesel production by reactive distillation

In the conventional biodiesel productivity (Zhang et al., 2003), there consists of reactor following by distillation. Transesterification is a reversible reaction which conversion is limited by the equilibrium. Using reactive distillation can be obtained higher conversion due to removing product continuously. In addition, combining unit may reduce capital and operating cost. That using reactive distillation integrating two units is a challenging task. The process is based on the conventional process (above process). The reactor and distillation column are replaced by the reactive distillation. In addition, most detail of process is similar to the conventional one. Figure 24 shows the biodiesel process using reactive distillation. There are three steps process which are reactive distillation, glycerin and alkali removing.

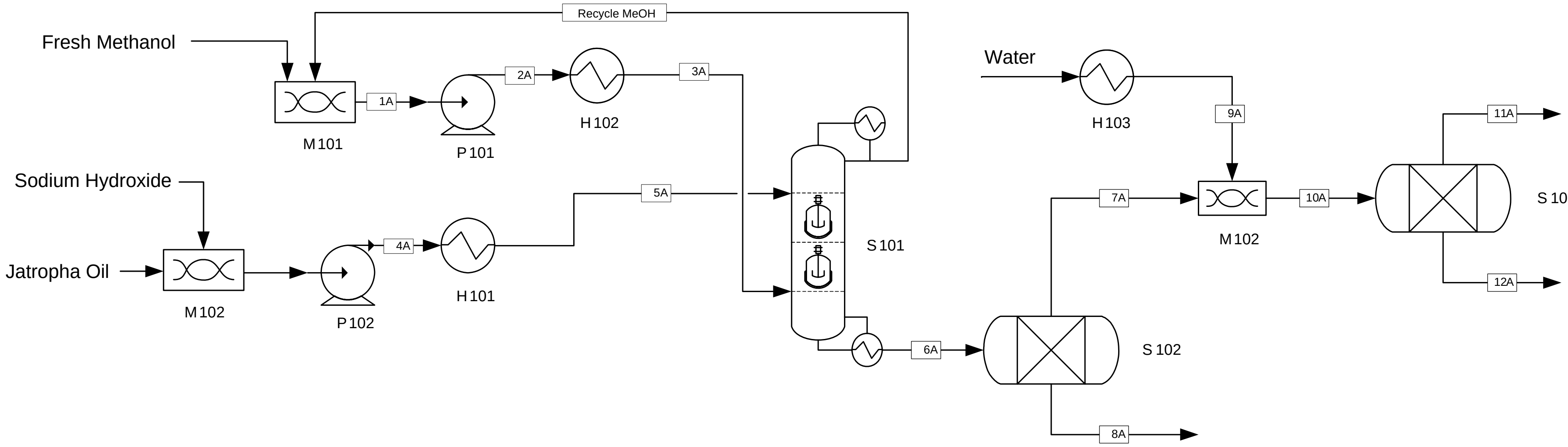


Figure 24 The biodiesel production by using reactive distillation

3.1 Reactive distillation

Biodiesel production from reactive distillation is similar to the conventional process. The transesterification and methanol recovery section were replaced by reactive distillation as seen in the Figure 25. This column were used both reaction and separation. Methanol 120 kg/h was mixed with recycle methanol in order to reduce fresh methanol. After that, it was heated to 60 ^{o}C which is an optimal condition for producing biodiesel and was sent to the reactive distillation. Jatropha oil 1,000 kg/h was mixed with sodium hydroxide, 1 % of Jatropha oil, and heated to 60 ^{o}C. Because catalyst can not evaporate in reactive distillation, Jatropha oil and sodium hydroxide should be fed at the top of column, while methanol should be fed at the lower of column. After mixing, it was charged to reactive distillation. The methanol and oil were feed at tray three. The transesterification reaction occurred in reactive distillation which the operating conditions are 1 reflux ratio, 0.6 boiler, 1 atm and 5 stages. The reaction zones were set at tray three. Methanol which was a product in the distillate stream was recycled and combined with fresh methanol. Biodiesel and other component was removed in the bottom stream and sent it to the purification section.

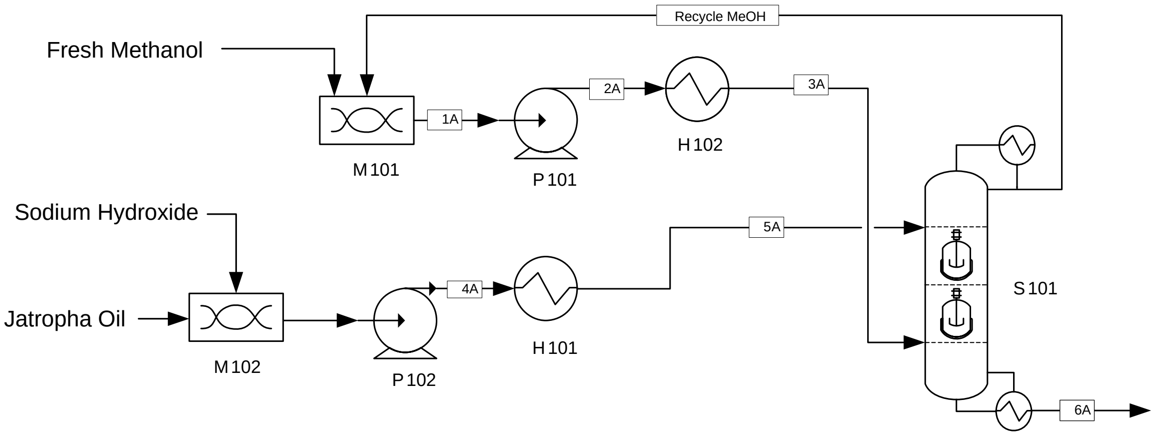


Figure 25 Reactive distillation section

3.2 Glycerin removal

The amount of glycerin can affect to the quality of biodiesel. The gravity unit (decanter) shown in the Figure 26 was introduced to separate for these propose. The glycerin phase which is lower than biodiesel was set as a second liquid and removed by the second liquid stream. The conditions in the decanter were set 25 ^{o}C and 1 atm. Biodiesel and other component was removed by first liquid stream and was sent to the alkali removing. The amount of glycerin in the biodiesel is 0.001% that was in standard (ASTM6751).

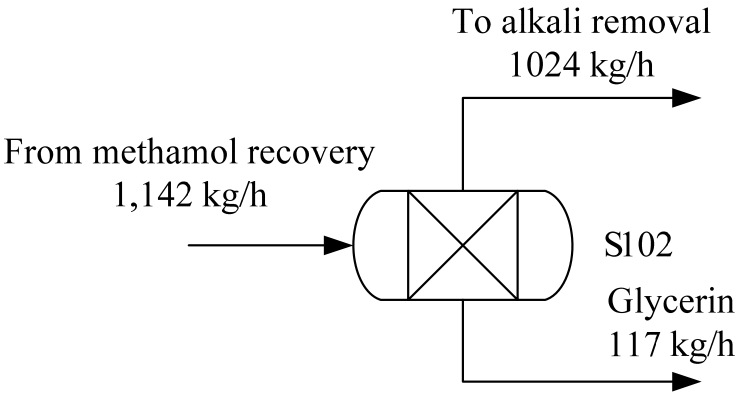


Figure 26 Glycerin removal section

3.3 Alkali removal

After removing glycerol, there is a lot of base catalyst dissolving in biodiesel. Therefore, the catalyst was should removed by using hot water. After adding water, catalyst was dissolved in to the water phase. There are two layers which are biodiesel and water layer. Figure 27 shows the decanter units used to separate biodiesel and water. Hot water 900 kg/h at 50 °C was used in order to dissolve catalyst. After the alkali removing section, the amount of catalyst in the biodiesel was less than 0.0005 % which meets the standard (ASTM6751).

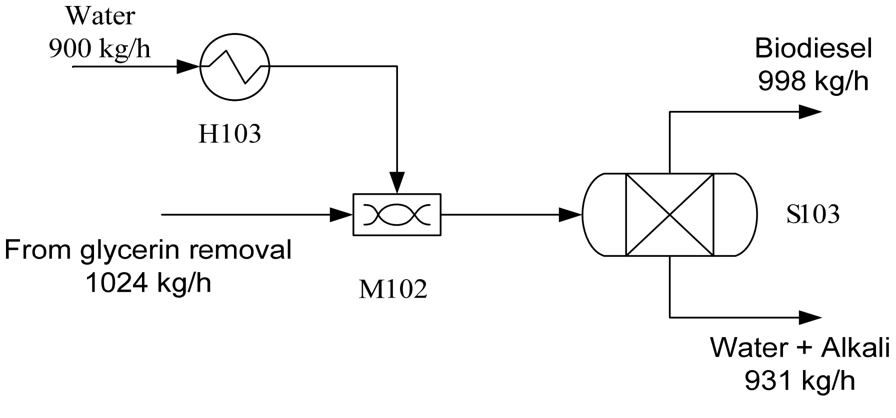


Figure 27 Alkali removal section

4. Base case of reactive distillation column

Due to the unavailable information of reactive distillation for biodiesel production, the base case details of column consisting five theoretical stages, 1 reflux ratio, 0.6 boilup ratio, 1 atm were assumed as a conventional process (Zhang, 2003). The reaction zones are set at stages three. Table 5 shows the streams table of column. In addition, the energy of condenser and reboiler are 0.089 Gcal/h and 0.045 Gcal/h, respectively.

Table 5 The streams table of column

Mass Flow kg/hr	Distillate	Bottom
TRIOLEIC	0	0.04551
TRI-LINOLEIC	0	0.03211
DI-OLEIC	0	0.05745
DI-LINOLEIC	0	0.142
MONO-OLEIC	0	0.00046
MONO-LINOLEIC	0	0.00038
METHYL-OLEIC	0	506.403
METHYL-LINOLEIC	0	405.894
GLYCEROL	0	80.293
METHANOL	192.629	0
NAOH	0	12
WATER	0	0

5. The comparison between conventional and reactive distillation

The procedure between two processes for biodiesel production which are transesterification, methanol recovery, glycerol and alkali removing are analogous. Reactive distillation integrating reaction and distillation in single was employed. The all parameters in the reactive distillation are set as the base case of reactive distillation. Figure 28 shows the comparison of biodiesel conversion in the bottom stream of distillation and reactive distillation. The biodiesel conversion in reactive distillation is higher than conventional about 1.6 %. The energy requirement of two processes was also compared as seen in the Figure 29. The reactive distillation process use energy more than conventional 1.36 %.

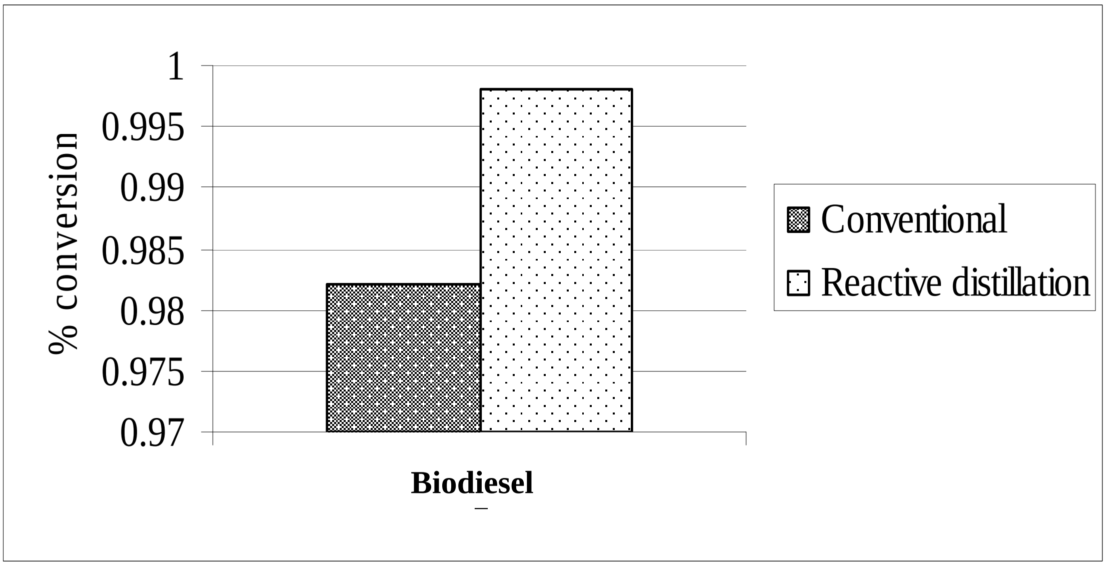


Figure 28 The comparison of conversion in the bottom stream between conventional and reactive distillation.

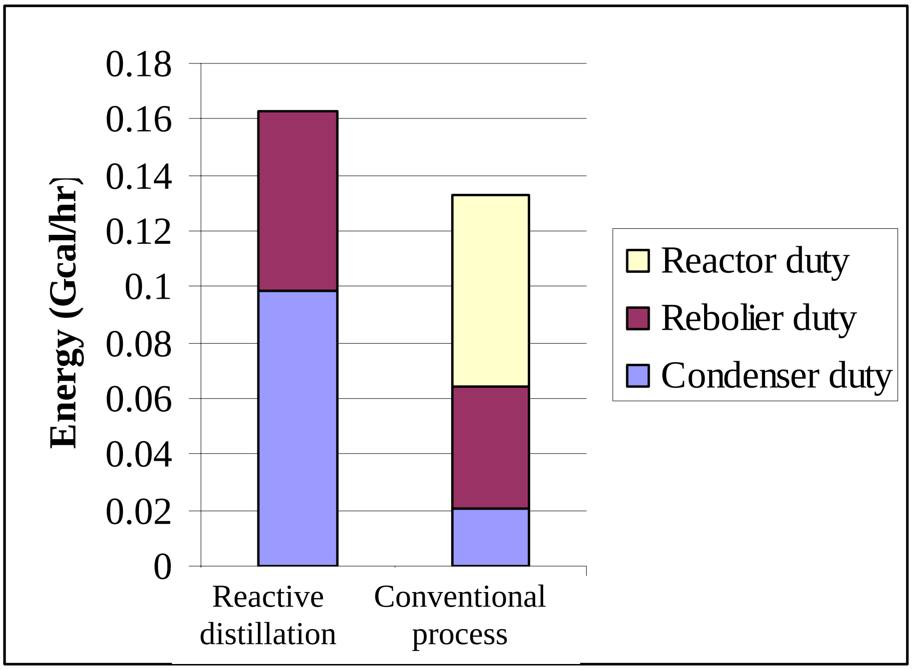
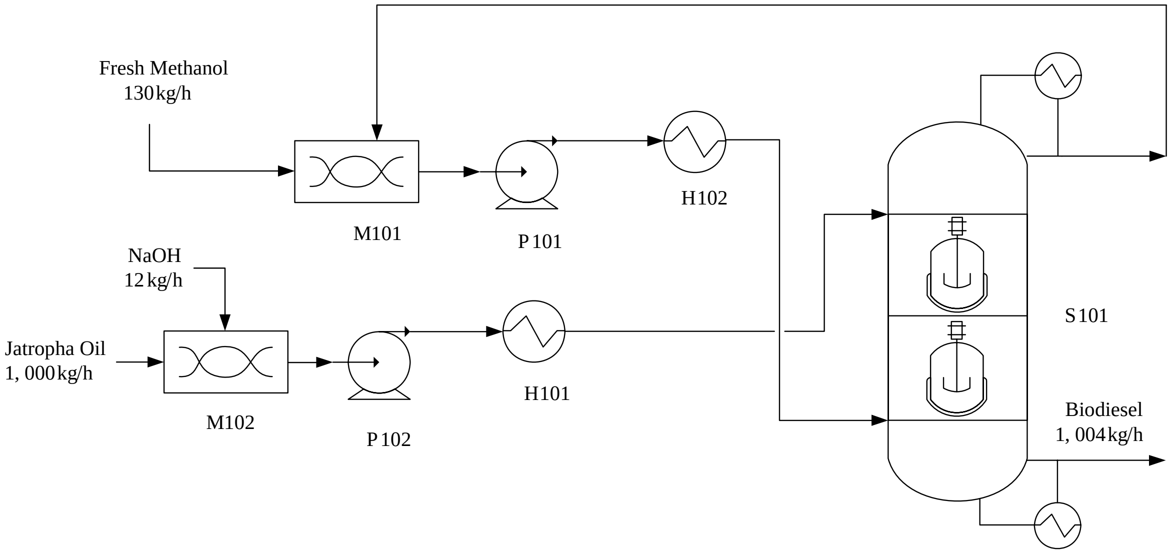


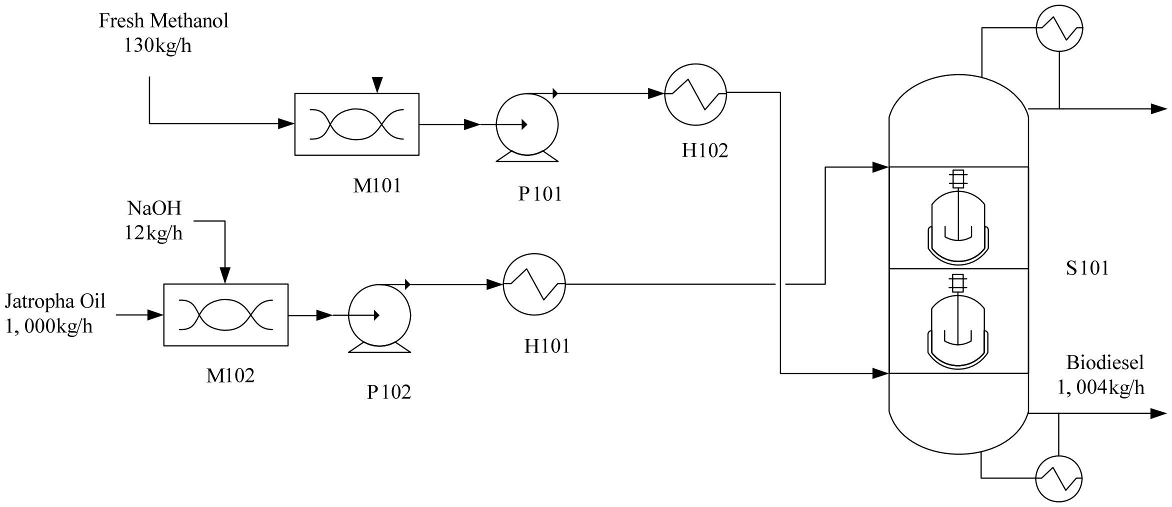
Figure 29 The comparison of energy consumption between reactive distillation and conventional process.

6. The comparison between with and without recycle stream

Because of high quantity methanol in feed stream, the methanol recovery unit was applied to recycle the excess methanol. The comparison between with and without recycle methanol in the reactive distillation system as seen in Figure 30 was investigated in this research. The amount of biodiesel in the recycle case was 1004 kg/h. In order to obtain similar value, the methanol inlet of without recycle should be 700 kg/h. Therefore, recycle methanol in the reactive distillation play an important role to reduce fresh methanol.



(a)



(b)

Figure 30 The reactive distillation (a) with recycle (b) without recycle

7. Optimum parameters of reactive distillation

Due to the limitations and constraints in the experimental, some of the important design parameters such as column pressure, reflux ratio, number of rectifying, number of stripping and number of reaction zones could not be studied experimentally. Therefore, the effect of parameter on the conversion of biodiesel needs to be evaluated by simulation. The simulations were based on base case of reactive distillation.

7.1 Height of reaction zone

Normally, height of reaction zone depends on the type of the catalyst. In the heterogeneous, height of reaction zone depends on the height of catalyst. On the other hand, in the homogeneous catalyst, it depends on the contacting between reactant in the column. In this research, homogeneous catalyst which is sodium hydroxide was used to increase the biodiesel conversion. Therefore, height of reaction zone was depended on the contacting between oil and methanol. The arrangements of two feed reactants were analyzed in order to find the optimum reaction zone. The light reactant feed should place above the heavy reactant (Cheng and Yu, 2005). Thus, methanol which is light reactant should be fed at the lower and oil should be fed at the top of column. The column conditions were based on base case. Only number of reaction zone was varied. Figure 31 demonstrates the number of reaction zone. According to the Figure 31, the number of reaction zone increases from one to six. The effect of reaction zone on the biodiesel conversion was illustrated in the Figure 32. It seems from the figure that the conversion increases as the reaction zone increase because the residence time in one stage is not sufficient. The increasing reaction zone after 2 stages does not effect on the conversion. Therefore, the optimum reaction zone is 2 stages.

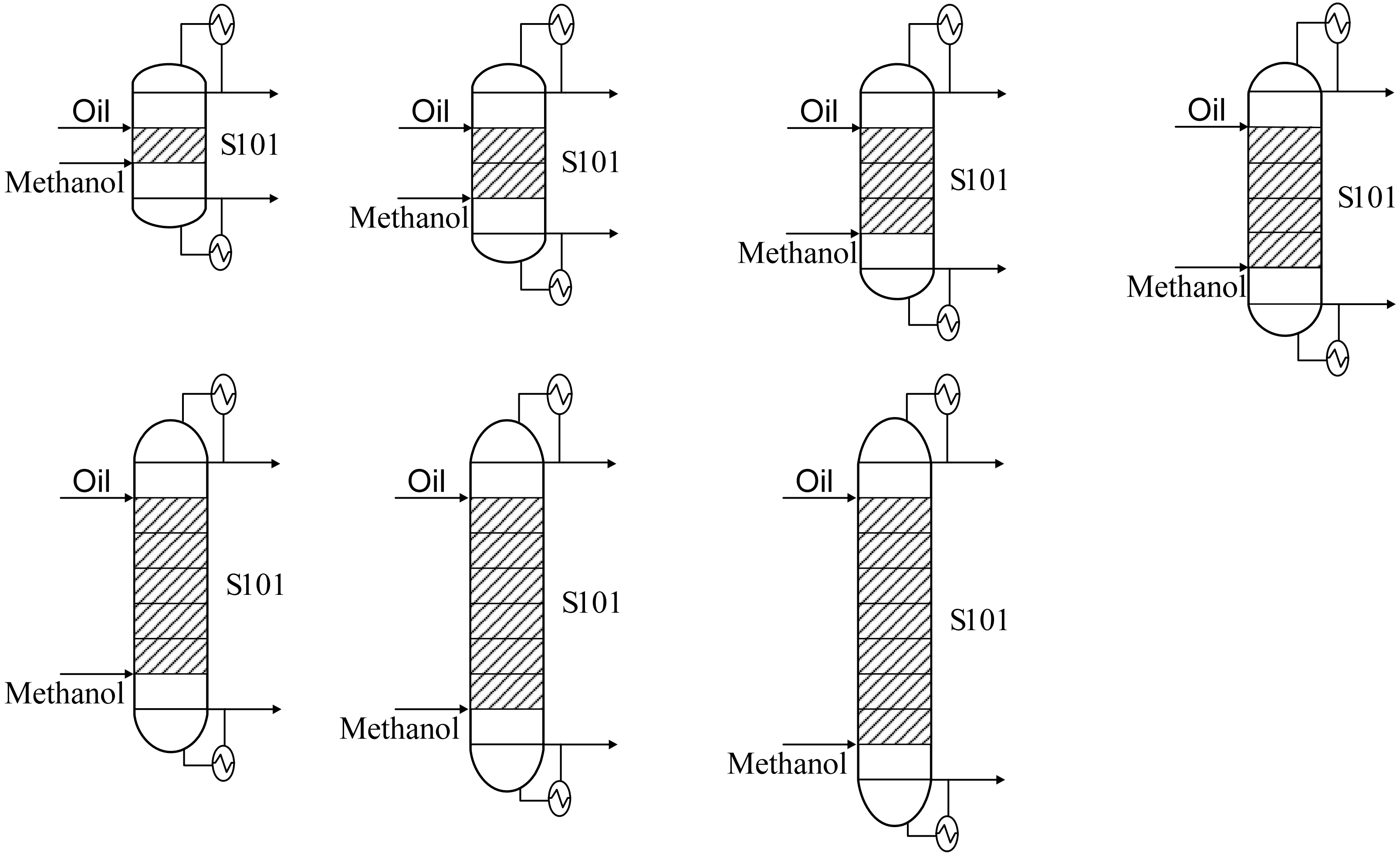


Figure 31 The number of reaction zone

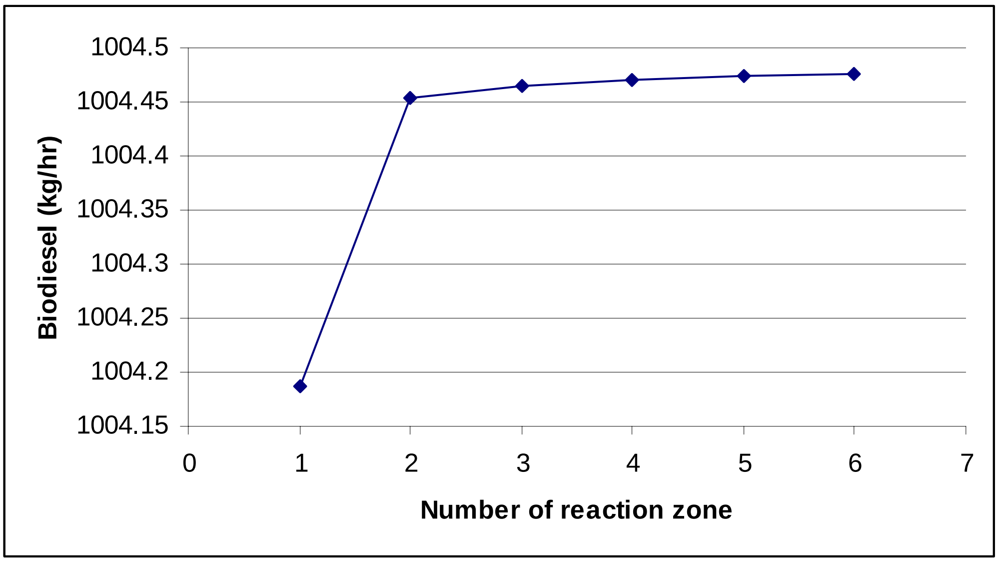


Figure 32 Effect of number of reaction zone on the biodiesel

7.2 Height of rectifying

Normally, the rectifying of distillation is used for removing the light component from the column. Higher amount of light component can be obtained by increasing the number of rectifying zone. However, in the reactive distillation, the rectifying zone should (a) prevent the loss of reactant (b) remove the product for increasing the conversion. Therefore, the effect of the number of rectifying zone on the biodiesel conversion was investigated. Figure 33 demonstrates the number of rectifying zone. The parameters in column were fixed as a base case. Only the number of rectifying is varied from one to four. The effect of the rectifying number was shown in the Figure 34. In this research, methanol which is a light component compared with other was separated by the top of column. The increasing rectifying zone can increase the amount of methanol in the distillate stream. The biodiesel decreased by this action because the methanol was removed from the column. Yield of biodiesel is not difference as increasing rectifying zone, while the amount of methanol increases. This methanol affects on the power requirement of pump. Higher methanol requires more power. Therefore, the best number of rectifying zone is 0.

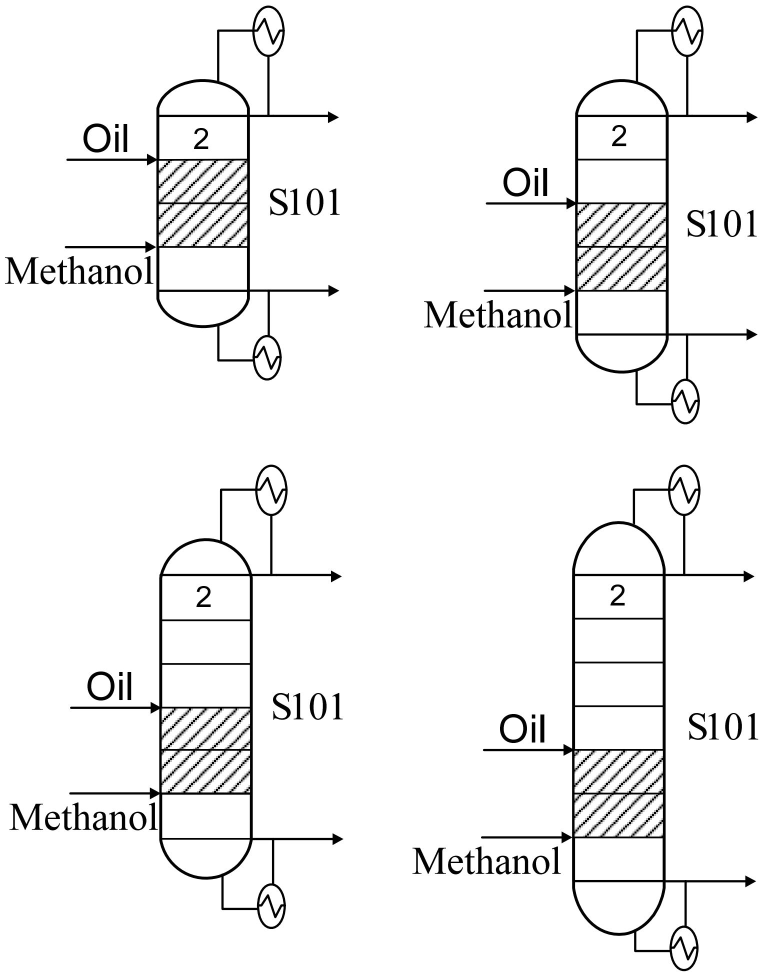


Figure 33 The number of rectifying.

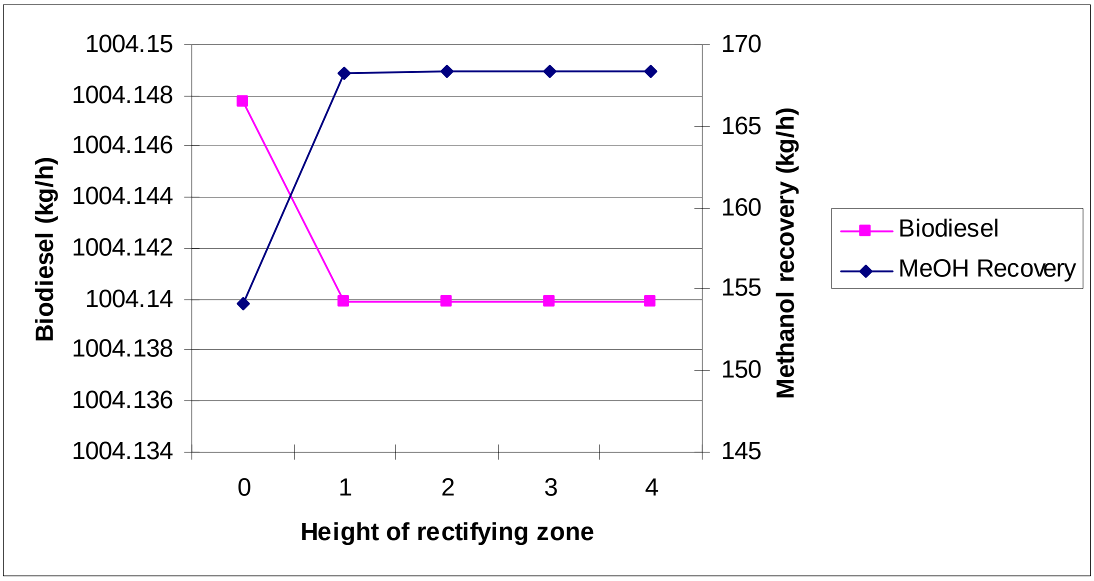


Figure 34 The effect of number of rectifying on the biodiesel

7.3 Height of stripping

In the reactive distillation, the stripping zone should (a) remove the product from the reaction zone (b) purify the product. However, the stripping zone of this process should move the products which are biodiesel and glycerol from the reaction zone. The conditions of column were fixed as base case. Only stripping zone was varied. Figure 35 demonstrates the number stripping zone. The stripping zone was varied from one to four. The effect of the number of stripping on the conversion was illustrated in the Figure 36.

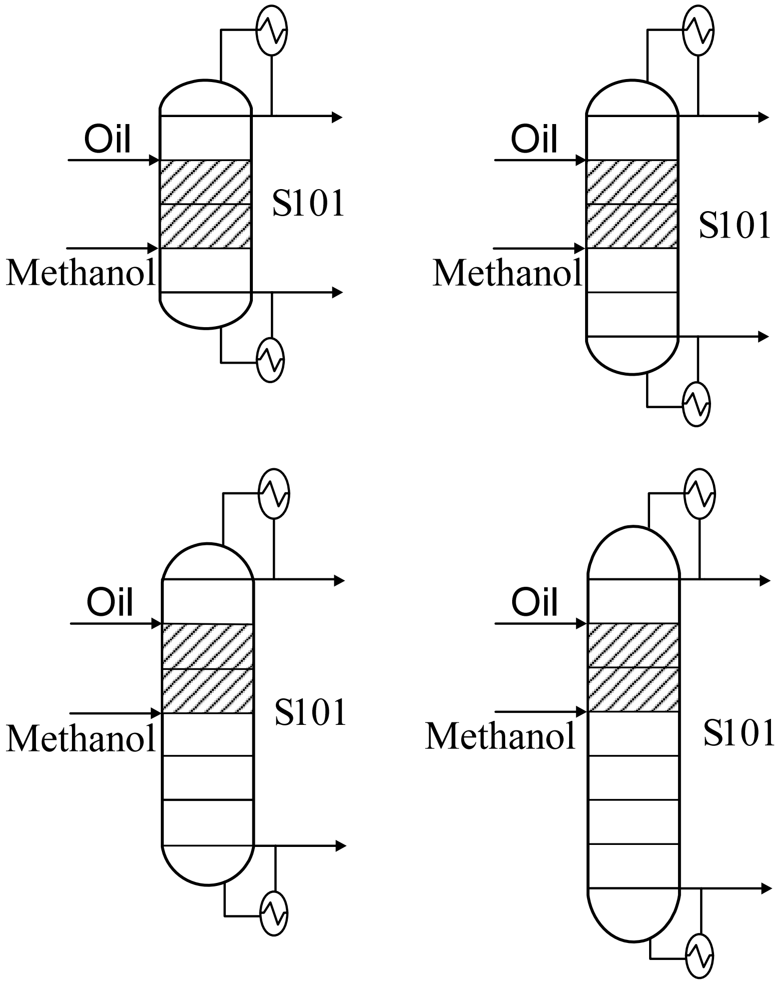


Figure 35 The number of stripping zone.

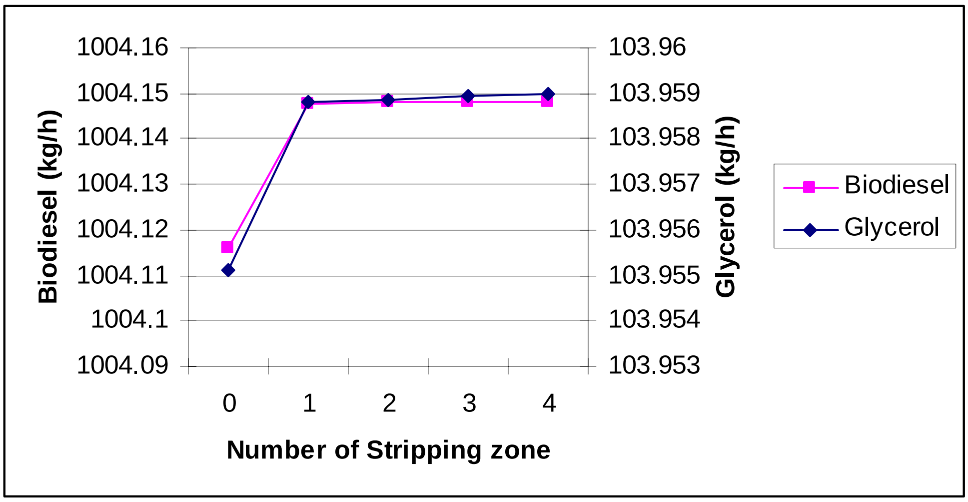


Figure 36 The effect of the number stripping zone on the biodiesel

As seen in the Figure 36, the amount of biodiesel and glycerol increases when stripping zone increases from one to two. Because increasing stripping zone, the products were removed higher. Therefore, the reaction is shift to right hand side, the product is higher. After two stripping, the product is still the same because of the limit of the separation. The yield of biodiesel does not change as increase number of stripping zone. Therefore, the optimum stripping should be zero.

7.4 Column pressure

Normally, the pressure of column is operated an economic ratio of heat-transfer costs and the improving separate (via increasing relative volatility with reducing pressure) (Kister, 1992). However, in the reactive distillation, the selection operating pressure has an effect on the reaction rate and the reaction equilibrium. The operation at maximum possible pressure is beneficial from the chemical equilibrium and rate of reaction consideration (Subawalla and Fair 1999, Bhatia *et al.* 2006). The parameters of column were fixed as base case. Only column pressure was varied. Figure 37 demonstrates the effect of column pressure on the biodiesel conversion and reboiler duty. The increasing operating pressure led to the reactive zone temperature and usually increased the reaction rate. According to the Figure 37, the biodiesel conversion is increased about 0.03 % when pressure increases from 1 to 4 atm while the reboiler duty increases about 33.35 %. From the results, the biodiesel slightly increases when pressure column is increased but the cost of column is considerable higher (Stitt, 2002). Therefore, operating at atmosphere pressure is satisfactorily.

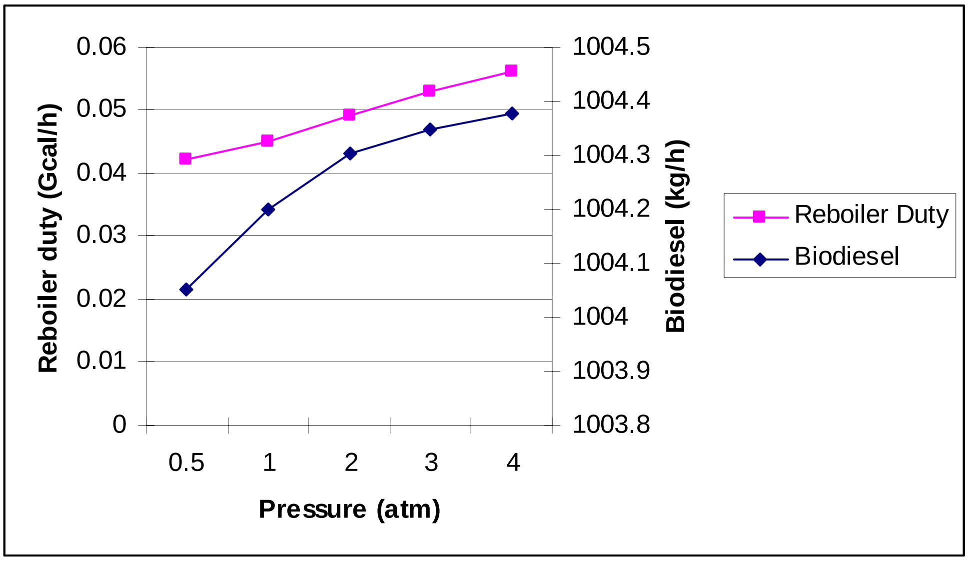


Figure 37 The effect of column pressure on the biodiesel and the reboiler duty.

7.5 Reflux ratio

The reflux rate and reflux ratio are important parameters and affect both reaction and separation performance in a reactive distillation column. The reflux ratio also significantly affects the reactive zone and residence time. The excess reflux ratio leads to the operating problem and insufficient residence time (Agreda *et al*., 1990; Chopade and Shubham 1997; Luo and Xiao 2001; Bhatia *et al*., 2006). The conditions of reactive distillation were fixed as base case. Only reflux ratio was varied from 0.1 to 60. Figure 38 demonstrates the effect of reflux ratio on the biodiesel conversion. An increase reflux ratio caused the product dropped. The biodiesel conversion increased by 0.011 % as the reflux ratio increased from 0.1 to 5. The biodiesel conversion did not change after 5 reflux ratio. Moreover, reflux ratio not only affects to conversion but also affects to the quantity of methanol in distillate stream. When the reflux ratio was increased, the methanol in the distillate stream was decreased. The amount of recycle methanol affect to the auxiliary equipment such as pump. If the amount of recycle is high, more power should be required to carry the methanol. Figure 39 displays the effect of reflux ratio on the amount of methanol in the overhead. The methanol recycle reduced 90 % when increased reflux ratio from 0.1 to 5. The methanol doesn't change after ten reflux ratio. An increase in reflux ratio will also increase the vapor rate within the tower, thus increasing expenses the condensers cost (Douglas, 1988). Figure 40 shows the effect of reflux ratio on the condenser duty. As the reflux ratio increased, the condenser increased too. After 5 reflux ratio, the condenser duty increased a little. Considering all of effect (the biodiesel conversion, the methanol recovery and condenser duty), the reflux ratio should be operated at ten.

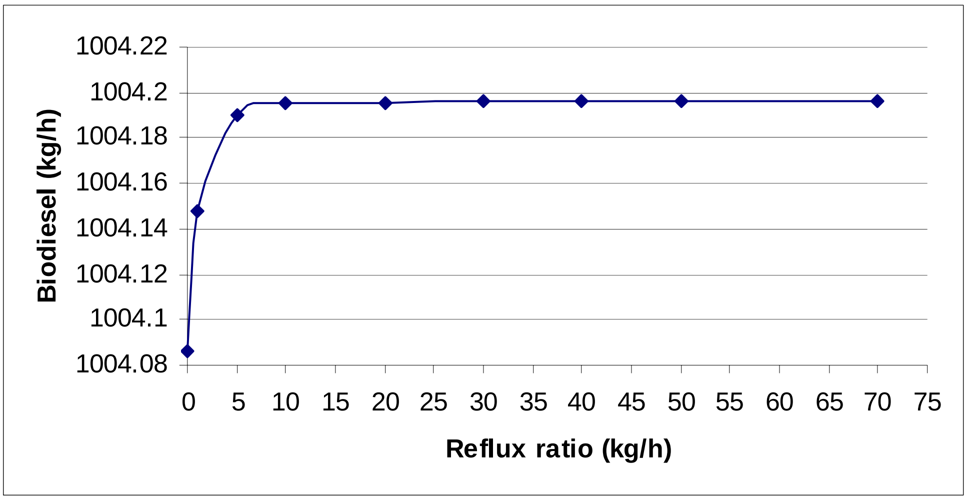


Figure 38 The effect of reflux ratio on the biodiesel

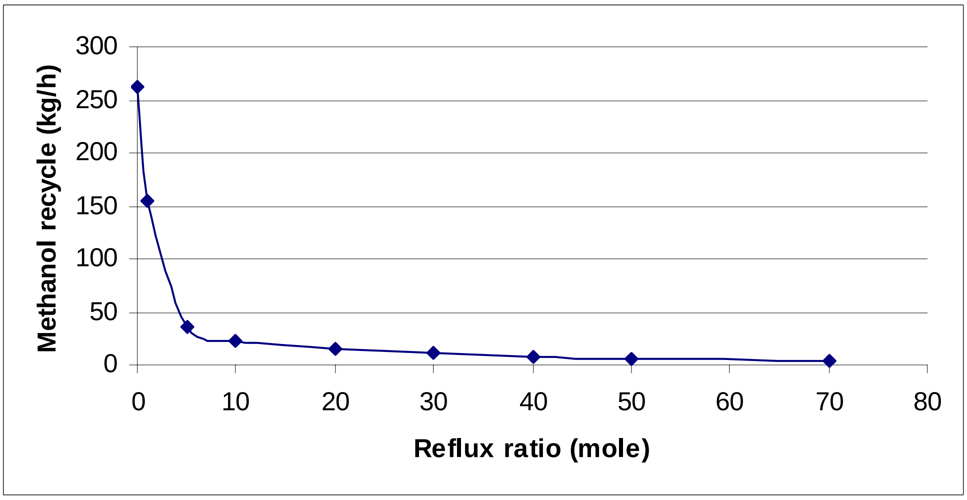


Figure 39 The effect of reflux ratio on the methanol recycle

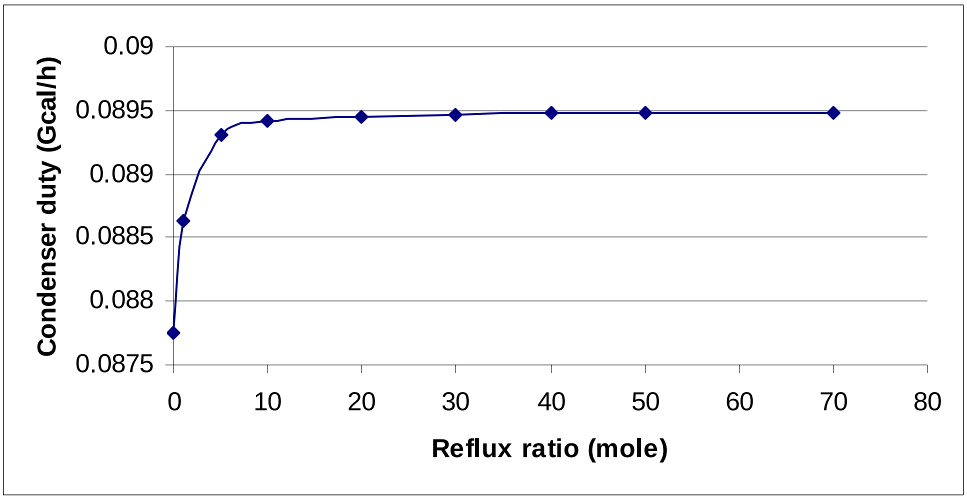


Figure 40 The effect of reflux ratio on the condenser duty

8. Dynamic simulation

Reactive distillation column has become an attractive unit operation in recent years due to its ability to reduce energy consumption and the number of equipment unit, even though, one of the main disadvantages is the nonlinear characteristics such as multiple steady states and high sensitivity to operating variables due to the coupling between separation and chemical reaction (Wang and Wong, 2006). Therefore, the controllability of reactive distillation production of biodiesel need to be investigated. The simulation results of biodiesel process were exported from ASPEN PLUS to ASPEN DYNAMIC for studying the performance of controller.

8.1 Feed oil disturbance

The effect of oil feed was studied on quality and quantity of biodiesel. Figure 41 demonstrates the biodiesel process which consists of control structure from default of ASPEN DYNAMIC. The regulatory control structure comprised of seven loops; 1) pressure control in the column 2) liquid level on top of column 3) liquid level on bottom of column 4) first liquid level of glycerin removal 5) second liquid level of glycerin removal 6) first liquid level of alkali removal 7) second liquid level of alkali removal.

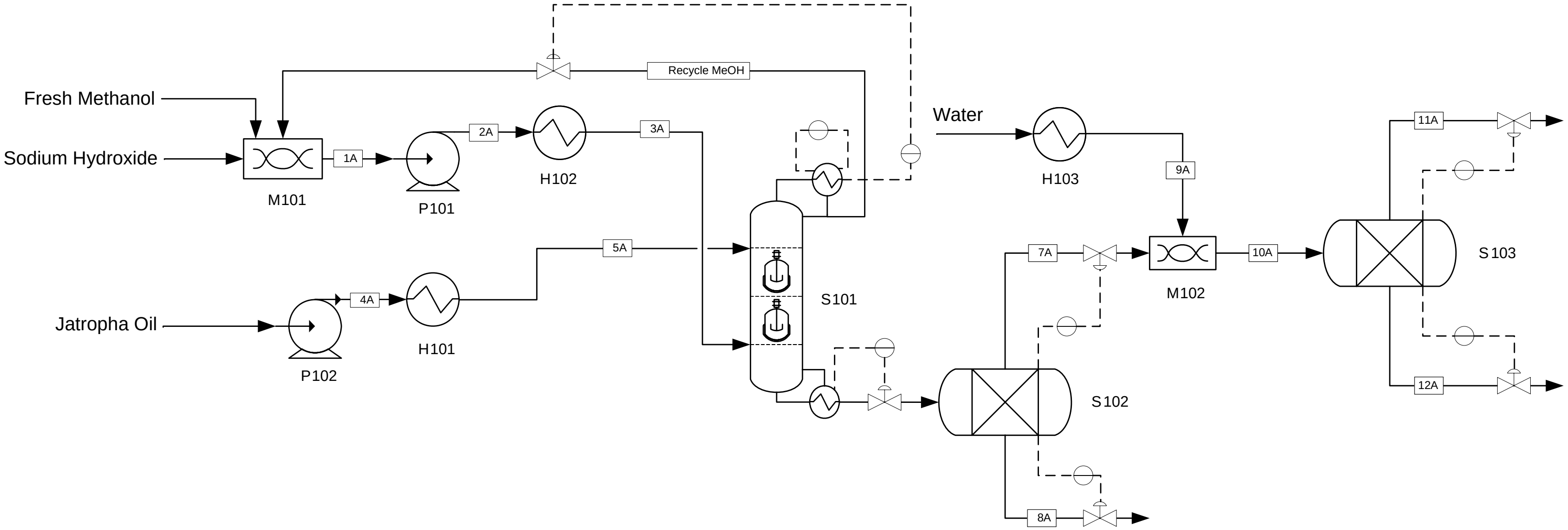


Figure 41 The Control structure of biodiesel process.

In order to control this process, PI controller was applied to control. However, some of parameter for control is not available. There are two techniques which are quarter amplitude decay and critical damped decay for finding this parameter. The quarter amplitude decay is suitable for this research because oil feed disturbance is set as step change. In addition, the Ziegler Nichols technique used to find parameter was showed in the Appendix B. Table 7 shows the parameters turning, K and $\tau_{i,}$, of each controller in case of increase and decrease oil feed.

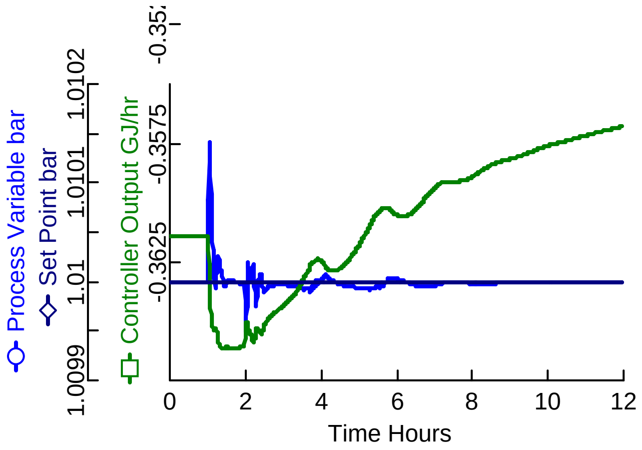
Table 7 The gain and integral time of each controller in case of increase and decrease oil feed by Ziegler Nichols technique

	Increase flow		Decrease flow	
	Proportional Gain	τ_I (min)	Proportional Gain	τ_I (min)
Pressure at Stages 1	45	0.5	65	1
Liquid level at stages 1	60	0.1	70	0.1
Liquid level at stages 5	55	0.5	60	0.7
Liquid 1 level of glycerin removal	50	1	45	1
Liquid 2 level of glycerin removal	50	0.1	50	0.1
Liquid 1 level of alkali removal	65	1	60	1
Liquid 2 level of alkali removal	60	1	70	1

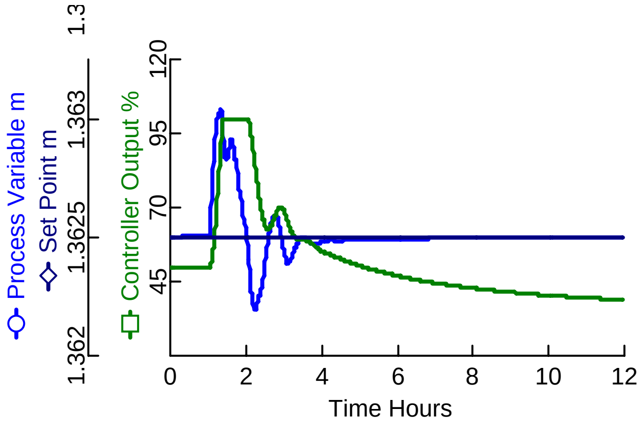
The Jatropha oil feed was then introduced as a disturbance. The feed oil was varied ± 2 % to 1,020 kg/h and 980 kg/h after 1 hour. The controller responses of increasing flow rate were demonstrated in the Figures 42, 43 and 44. As can see in the Figure 42, the controller response of column was changed after disturbance the oil feed at 1 hour. All controller of column can be controlled the column pressure, liquid level at stage 1 and liquid level at stage 5 to desired set point. Figure 43 shows the controller response of decanter for removal glycerin, two level control of decanter can be controlled the liquid level to desire set point. Figure 44 demonstrates the controller response of decanter for removing alkali, the firstly and second liquid level controller can be controlled the liquid level to desire set point. Therefore, all default controller can be controlled this process in case of increasing oil feed + 2%. In case of decreasing oil feed – 2%, on the other hand, the responses of decreasing flow rate were demonstrated in the Figures 45, 46 and 47. Figure

45 displays the controller response of column. The controller can be controlled the column pressure, liquid level at stage 1 and liquid level at stage 5 to desired set point. Figure 46 shows the controller response of decanter used to remove glycerin. The two level controller of decanter can be controlled the liquid level to desire set point. Figure 47 demonstrates the controller response of decanter for removing alkali, the firstly and second liquid level controller can be controlled the liquid level to desire set point. The default controller can control system in case of decreasing oil feed.

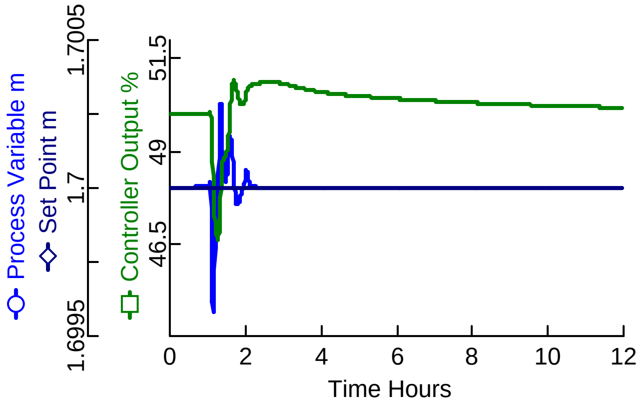
As seen from the figure, all controller of process can control system in case of increasing and decreasing oil feed. Figures 48 and 49 illustrate the effect of feed oil ± 2 % from 1000 kg/h on quantity and quality, respectively. In case of increasing feed oil, amount of biodiesel was increased from 995kg/h to 1,012 kg/h, and the biodiesel still meet on specification of EN and ASTM standard. The decreasing of oil feed, the amount of biodiesel was also decreased from 995 kg/h to 977 kg/h. The controller in case of both increase and decrease oil feed need 7.5 hour to drive the quantity of biodiesel output to steady state. This control system can work properly in this simulation (± 2 % oil feed) due to the capacity of system which is fixed as steady state. However, in case of high disturbance, the capacity of system should be changed to fit with load.



(a)

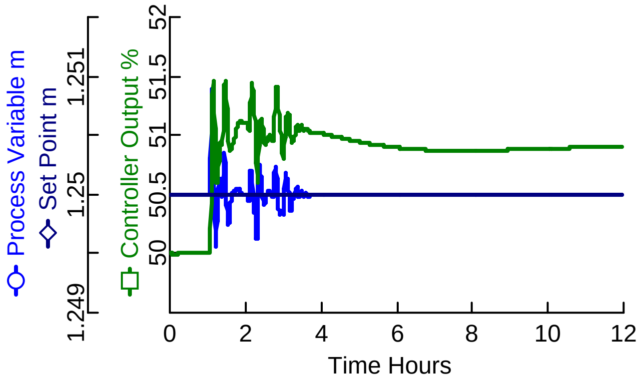


(b)

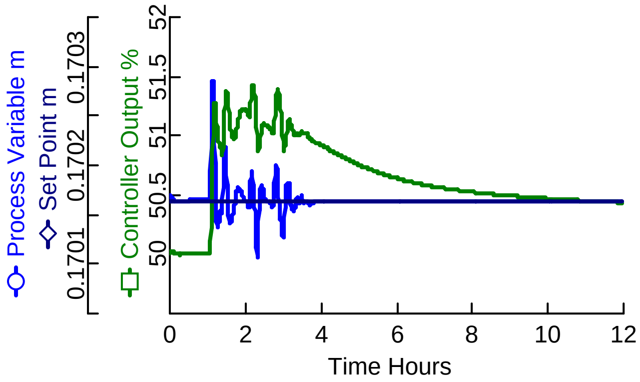


(c)

Figure 42 The controller response of column in case increasing oil feed (a) column pressure (b) liquid level on top of column (c) liquid level on bottom of column



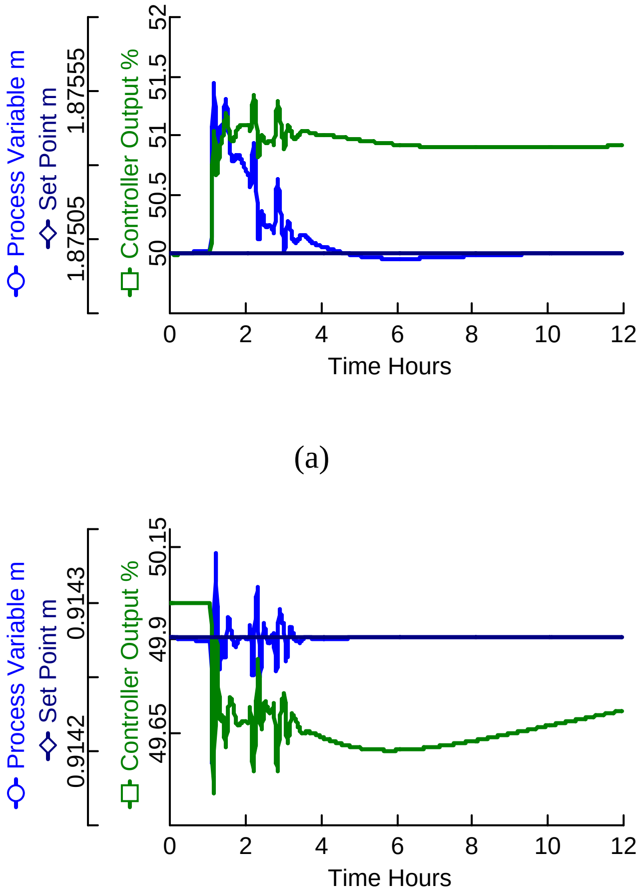
(a)



(b)

<u>Figure 43</u> The controller response of glycerin removal in case of increasing oil feed

(a) first liquid level of glycerin removal (b) second liquid level of glycerin removal



(a)

(b)

Figure 44 The controller response of alkali removal in case of increasing oil feed (a) first liquid level of alkali removal (b) second liquid level of alkali removal.

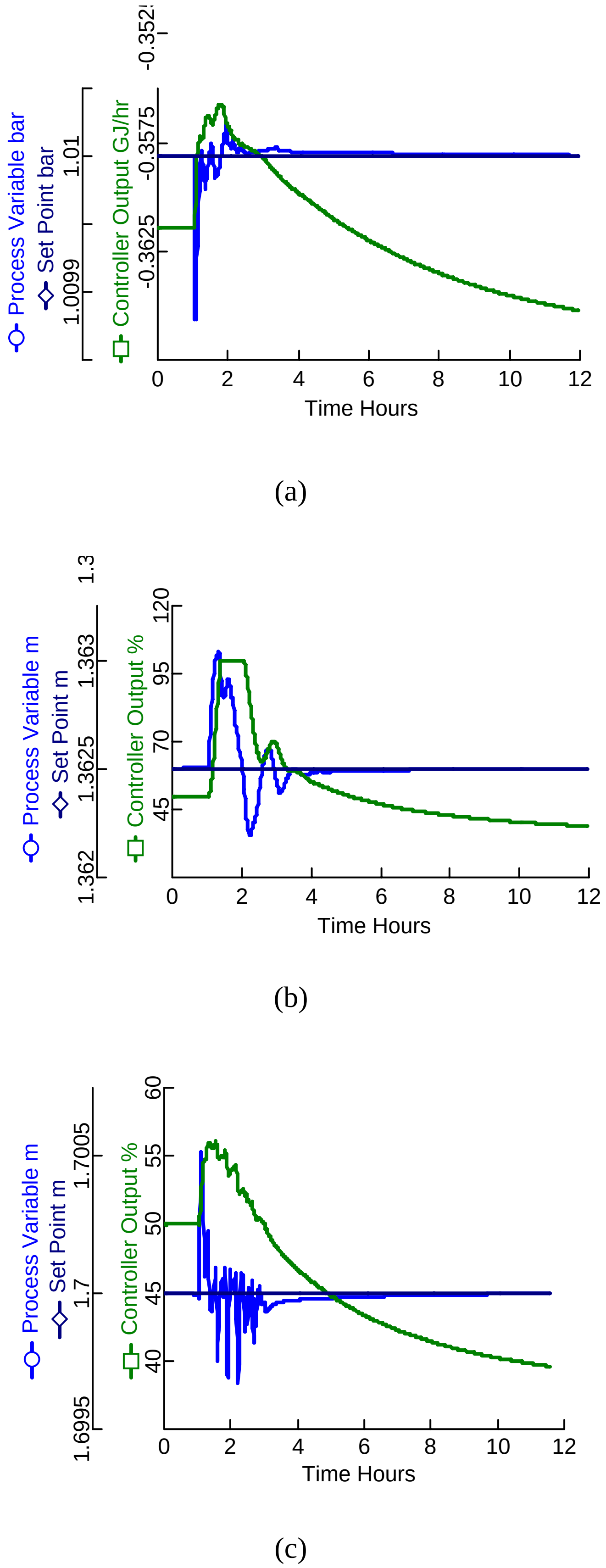
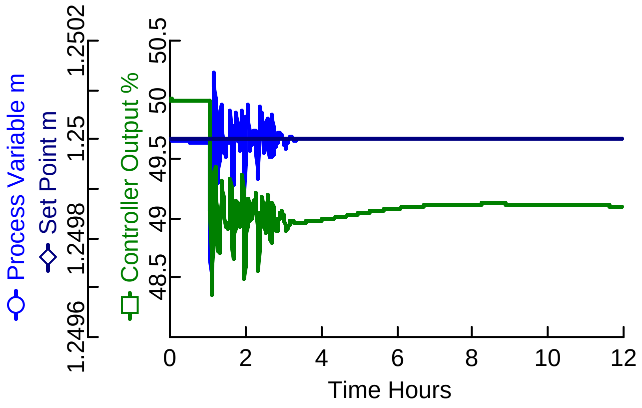
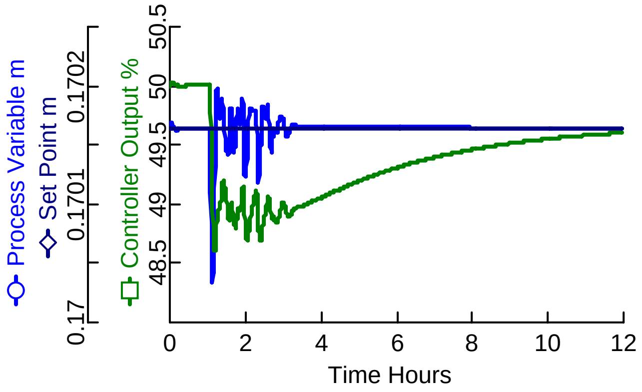


Figure 45 The controller response of column in case of decreasing oil feed (a) column pressure (b) liquid level on top of column (c) liquid level on bottom of column

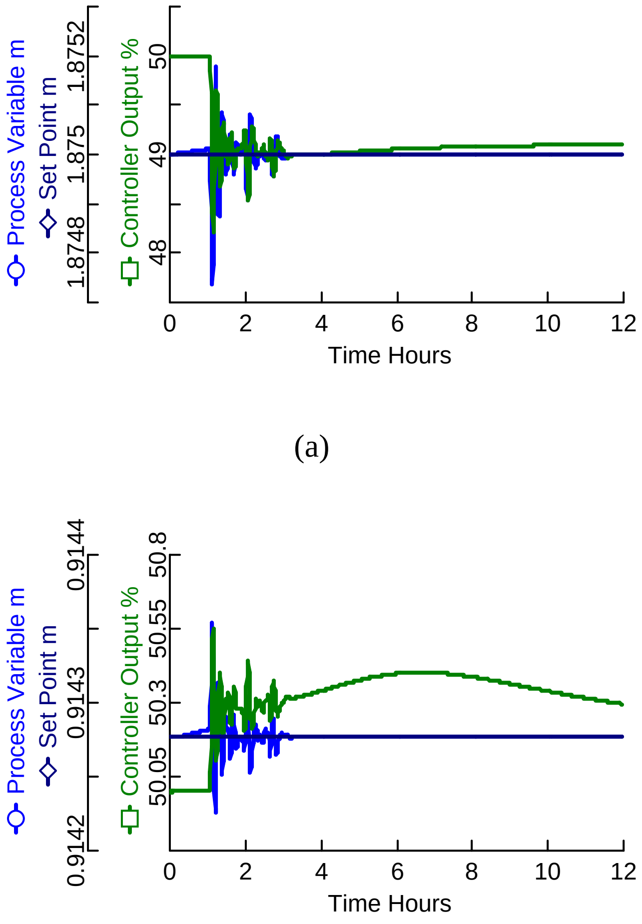


(a)



(b)

Figure 46 The controller response of glycerin removal in case of decreasing oil feed (a) first liquid level of glycerin removal (b) second liquid level of glycerin removal

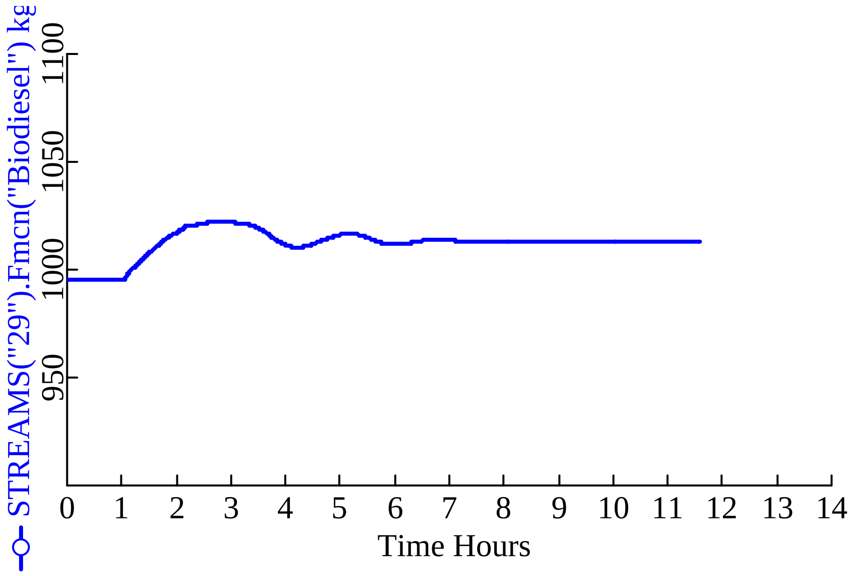


(a)

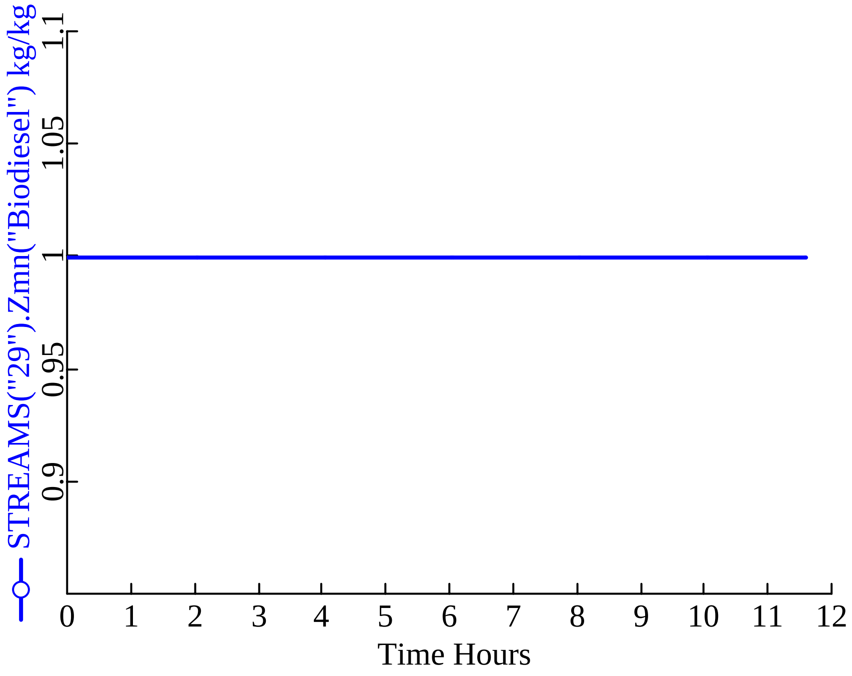
(b)

Figure 47 The controller response of alkali removal in case of decreasing oil feed

(a) first liquid level of alkali removal (b) second liquid level of alkali removal.

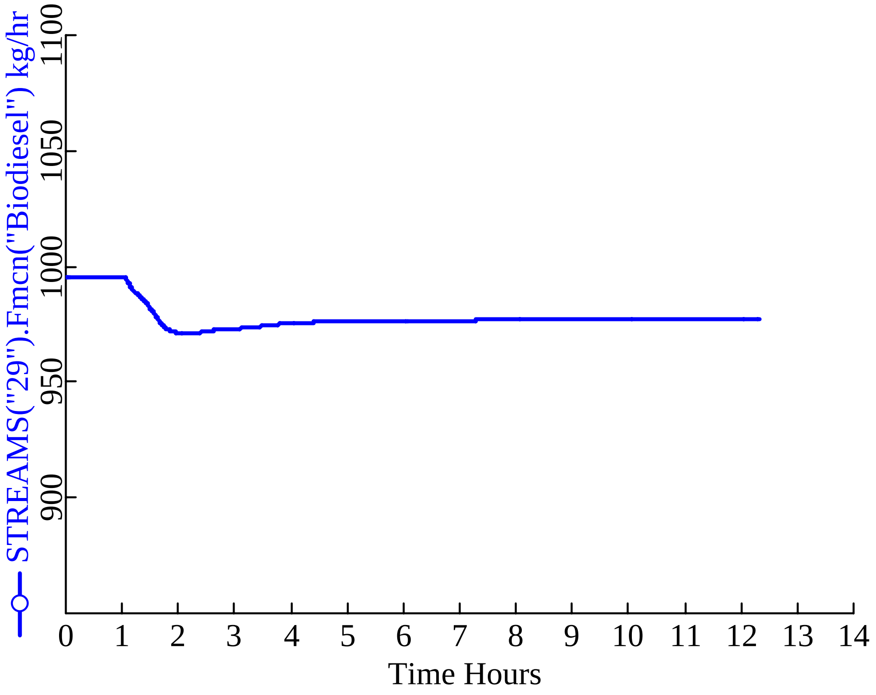


(a)

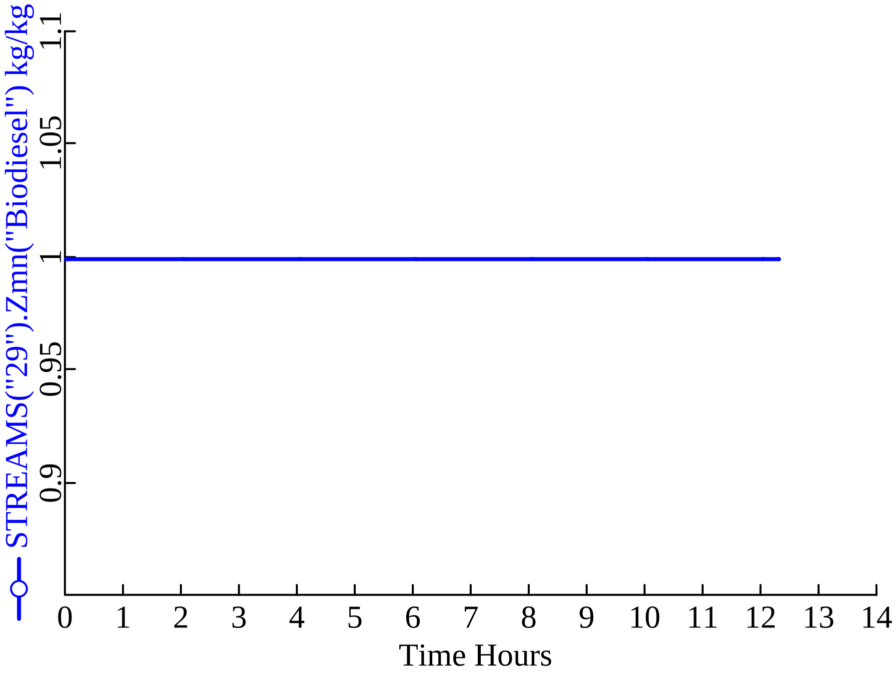


(b)

Figure 48 The effect of changing feed oil from 1,000 kg/h to 1,020 kg/h on (a)quantity of biodiesel b) quality of biodiesel



(a)



(b)

Figure 49 The effect of changing feed oil from 1,000 kg/h to 980 kg/h on (a) quantity of biodiesel b) quality of biodiesel