

## Design and test of a continuous reactor for palm oil transesterification

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### Abstract

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The continuous reactor for transesterification of refined palm oil with methanol was designed and tested. The reaction condition was focused at ambient pressure, temperature of 60°C, molar ratio of alcohol to oil of 6:1, and NaOH of 1.0 %wt of oil. The designed reactor was in a form of a 6-stage mechanically stirred tank. Rushton turbines, with 4 standard baffles, and plates with a small opening were installed inside. The reactor has a simple form which could be conveniently constructed and operated. The reactor could produce methyl esters (ME) with purities ranging from 97.5-99.2 %wt within residence times of 6-12 minutes in which its production performance was equivalent to a plug flow reactor and the power consumption of a stirrer in the range of 0.2-0.6 kW/m<sup>3</sup> was required. The reaction modeling based on a homogeneous concentration field with reaction kinetics could accurately predict the produced purities of ME. The production yields by weight of final product and of ME to the fed oil were 94.7 and 92.3%, respectively. The developed continuous reactor has good potential for producing ME to be used as biodiesel.

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**Key words** : biodiesel, continuous reactor, methyl esters, refined palm oil, transesterification

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## บทคัดย่อ

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การออกแบบและทดสอบถังปฏิกรณ์แบบต่อเนื่องสำหรับการทำปฏิกิริยา  
ทรานเอสเตอริฟิเคชันของน้ำมันปาล์ม

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งานวิจัยนี้ได้ออกแบบและทดสอบถังปฏิกรณ์แบบต่อเนื่องสำหรับการทำปฏิกิริยาทรานเอสเตอริฟิเคชันของน้ำมันปาล์มกับเมทานอล โดยได้กำหนดสภาวะการทำปฏิกิริยาไว้ที่ความดันบรรยากาศ อุณหภูมิ 60°C อัตราส่วนโดยโมลของแอลกอฮอล์ต่อน้ำมัน 6:1 และความเข้มข้นของโซเดียมไฮดรอกไซด์ 1.0% โดยน้ำหนักน้ำมัน ถังปฏิกรณ์ที่ออกแบบมีลักษณะอย่างง่ายซึ่งสามารถสร้างและใช้งานได้อย่างสะดวก โดยเป็นถังกวนเชิงกล 6 ชั้น มีใบพัดกวนแบบ Rushton แผ่นพับเฟิลมาตรฐานจำนวน 4 แผ่น และแผ่นวงแหวนที่มีรูเปิดขนาดเล็กติดตั้งอยู่ภายใน ถังปฏิกรณ์ที่ออกแบบสามารถผลิตเมทิลเอสเตอริ์ได้ที่มีความบริสุทธิ์ 97.5-99.2% โดยน้ำหนัก ภายในระยะเวลาการทำปฏิกิริยา 6-12 นาที ซึ่งสมรรถนะการผลิตของถังสามารถเทียบเท่ากับถังปฏิกรณ์แบบ Plug Flow Reactor และต้องการกำลังเพื่อจ่ายให้กับใบพัดกวนในช่วง 0.2-0.6 kW/m<sup>3</sup> นอกจากนี้การจำลองแบบปฏิกิริยาบนพื้นฐานของสนามความเข้มข้นเอกซิทและจลนศาสตร์ของปฏิกิริยาสามารถทำนายความบริสุทธิ์ของเมทิลเอสเตอริ์ที่ผลิตได้อย่างแม่นยำ ผลได้โดยน้ำหนักของผลิตภัณฑ์สุดท้ายและของเมทิลเอสเตอริ์ต่อน้ำมันปาล์มเริ่มต้นเท่ากับ 94.7 และ 92.3% ตามลำดับ ถังปฏิกรณ์แบบต่อเนื่องที่ถูกพัฒนานี้มีศักยภาพที่ดีสำหรับการผลิตเมทิลเอสเตอริ์เพื่อใช้เป็นไบโอดีเซล

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Presently, biodiesel has become more attractive in many countries. In Thailand, the highest potential source for biodiesel production is oil from palm fruits. The success of commercially using biodiesel from palm oil is an important factor for Thailand's sustainable development. The positive economic effects and an increase of energy security would be achieved.

The most common method for biodiesel production is the batch process at temperature near the boiling point of alcohol. The typical molar ratio of alcohol to oil is 6:1. The catalyst used is NaOH or KOH at the concentration in the range of 0.1-1.0 %wt of oil. However, for commercial production, continuous processes can provide lower production cost and a more uniform product quality than batch processes (Ma and Hanna, 1999; Srivastava and Prasad, 2000; Nouredini *et al.*, 1998; Darnoko and Cheryan, 2000a).

A number of continuous reactors for biodiesel production have been developed. Darnoko

and Cheryan (2000a) evaluated a continuous stirred tank reactor (CSTR) for transesterification of refined palm oil (RPO). Nouredini *et al.* (1998) examined a continuous process consisting of 2 motionless mixers, a high-shear mixer, and a residence tube for transesterification of soybean oil. Harvey and Mackley (2002) assessed an oscillatory flow reactor for transesterification of rapeseed oil. According to the European Union standards for alternative diesel fuels, the minimum acceptable purity of fatty esters as saleable biodiesel is 96.5 %wt (Karaosmanoglu *et al.*, 1996). Simple and complex reactors could produce saleable biodiesel. However, the required residence time depends on reaction condition and residence time distribution (RTD) of a continuous reactor.

The important aim of this work was to develop an effective continuous reactor for transesterification of RPO with methanol to produce saleable biodiesel. The reaction condition was focused at ambient pressure, temperature of 60°C,

molar ratio of alcohol to oil of 6:1, and NaOH of 1.0 %wt of oil. The design objective was to obtain a simple reactor whose production performance is equivalent to a plug flow reactor (PFR). A test was performed to evaluate the production performance, the RTD, and optimum process parameters focusing on the stirrer speed and the power consumption of the stirrer.

### Design of a Continuous Reactor

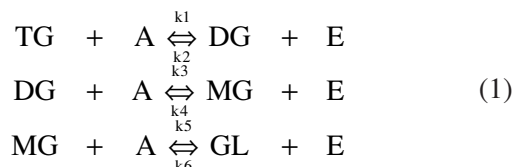
Three main steps were performed to design the continuous reactor. First, the kinetics of RPO transesterification in a batch reactor was studied to determine the intrinsic rate and appropriate reaction rate constants. Secondly, the RPO transesterification in continuous reactors was simulated to determine the optimum RTD and to predict the required residence time for producing saleable biodiesel. Finally, the effective process mixer was designed to be used as a continuous reactor.

#### 1. Kinetic Study of RPO Transesterification in a Batch Reactor

Darnoko and Cheryan, (2000b) studied the kinetics of RPO transesterification with methanol at ambient pressure and 6:1 molar ratio of alcohol to oil. They reported that the optimum temperature and KOH concentration were 60°C and 1.0 %wt of oil, respectively. The kinetic mechanisms satisfying the reaction were a pseudo-second order mechanism for an initial stage of the reaction followed by a first order or a zero order mechanism.

However, a second order mechanism without the shunt reaction as shown in the following equation (Noureddini and Zhu, 1997) was more convenient for predicting the reaction (Leevijit *et al.*, 2004a). Consequently, the experiment for transesterification of RPO with methanol in a well-mixed batch reactor was performed in our previous work (Leevijit *et al.*, 2004a). The reaction condition focused on the reported optimum condition, but NaOH was used instead of KOH. The intrinsic rate was obtained and the reaction rate constants ( $k_1 - k_6$ ) were determined to be  $1.057 \times 10^{-2}$ , 0.000,  $1.184 \times 10^{-1}$ ,  $8.187 \times 10^{-2}$ ,  $1.310 \times 10^{-1}$  and

$2.011 \times 10^{-3}$  l/mole.s. The overall goodness of fit to predict the purity of methyl esters (ME), the weight percentage of ME in the final product on a glycerol-free basis, during the reaction was quantified through 2 standards including the correlation coefficient ( $R^2$ ) and the mean relative deviation (MRD). They were 0.9936 and 1.835%, respectively.



where TG, DG, MG, GL, A and E are triglyceride, diglyceride, monoglyceride, glycerol, alcohol and esters, respectively.

#### 2. Simulation of RPO Transesterification in Continuous Reactors

The simulation study of RPO transesterification in continuous reactors was performed in our previous work (Leevijit *et al.*, 2004b) by using a classical reaction modeling method (a homogeneous concentration field with the obtained reaction kinetics). The system models of continuous reactors were constructed as they consisted of  $n$  tanks in a series of an equal residence time ideal CSTR. The simulations were run at various numbers of  $n$  and various residence times. The results revealed that, for producing saleable biodiesel, an optimum RTD was 6-ideal CSTRs in a series. The higher RTD did not give a significant decrease of the required residence time. The optimized reactor could produce ME at purities of 96.8-98.7 %wt within residence times of 3-12 minutes and its production performance was very close to a PFR.

#### 3. Process Mixer Design

The design of the mixer consisted of conceptual and detailed designs. The conceptual design was performed to obtain an appropriate configuration and the detailed design was performed to specify dimensions and to estimate operation parameters. Without specific reference, most of the information in this section refers to

Paul *et al.* (2004), Oldshue (1983), and Perry *et al.* (1997).

### 3.1 Conceptual design

A good understanding of process requirements was necessary to design an effective process mixer. The crucial process requirements of this work are listed in Table 1. Mainly, the reactor should have a simple form that could be conveniently constructed and operated. The simulation results revealed that, although the glycerol separation during the reaction to accelerate the reaction rate was not applied, the 6-ideal CSTRs in a series could produce saleable biodiesel within a short residence time. So, glycerol separation during the reaction was omitted in this design. The mixer would be applied for a liquid-liquid reaction. Based on the dynamic viscosities at 38°C of main fluids including RPO (36.4 cP) and ME (5.0 cP) (Srivastava and Prasad, 2000), the working range of the mixer could be roughly classified to be low-viscosity blending (< 5,000 cP).

For transesterification of vegetable oils, Nouredini and Zhu (1997) found that oil and alcohol formed a two-phase liquid system as initially introduced to the reactor so that intense mixing was required to boost the reaction rate. As ME was formed, it acted as a mutual solvent for the reactants. A single-phase system was then achieved and intense mixing became unnecessary. Nouredini *et al.* (1998) also found that the purity of ME in a continuous transesterification process decreased when excess mixing was supplied. In addition, to obtain the predicted production performance based on a homogeneous concentration field modeling, the reactor also had to supply sufficient mixing relative to flow rate and reaction rate to produce a homogeneous concentration field as modeled.

A number of process mixers could be applied for liquid-liquid blending. Either conventional mechanically stirred vessels or novel types of in-line mixers can be used. In this design, the former type was preferred for convenient construction and operation. Darnoko and Cheryan (2000a) measured the RTD of a CSTR made from a beaker and a magnetic stirrer for transesterific-

ation of RPO. The RTD followed an ideal CSTR pattern. This result revealed that it was not difficult to produce an ideal CSTR for transesterification of RPO. So, it was reasonable to design the desired continuous reactor as it consisted of 6 units of CSTR in a series. Furthermore, the arrangement in the form of vertically stacked type in which each unit was separated from each other by a plate with a small opening and all turbines were fixed on one shaft could provide the economic design.

Although there was no explicit information stating that a standard 6-blade-disk turbine (Rushton turbine) was the best choice for this application, its relations between shear rate and pumping capacity, and its flow patterns were very interesting. It generated both shear rate, which promotes a mixing intensity per revolution, and pumping capacity, which promotes a homogeneous concentration field. It also generated less undesirable axial flow along the shaft, which is the cause of inter-stage mixing of the multistage mixer. Hence, the Rushton turbine was selected. Four standard baffles at one-twelfth of the tank diameter were also added to promote the better mixing. The general configuration of the continuous reactor is shown in Figure 1.

### 3.2 Detailed design

The detailed design was performed to specify geometries, sizes, and operation parameters (the power consumption of a stirrer and the stirrer speed) of the reactor. In this work, the design was performed based on available information with the assumptions and constraints as shown in Table 1. The design condition was selected with the shortest predicted residence time that could produce saleable biodiesel.

**1) Assumptions:** Although the composition of reaction mixture and its physical properties changed along the degree of conversion of reaction, our simulation has shown that ME was the largest portion in the 6-ideal CSTRs in a series. So, the working fluid was assumed as ME to simplify the problem. Its available reported dynamic viscosity of 5.0 cP at 38°C and density of 0.880 kg/l (Srivastava and Prasad, 2000) were used

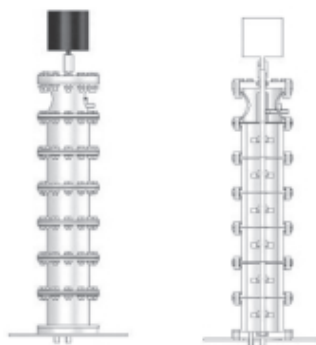


Figure 1. The configuration of a 6-stage continuous reactor.

Table 1. The process requirements, assumptions, and constraints for design the process mixer.

<b>Design reaction condition</b>	ambient pressure, temperature 60°C, molar ratio of alcohol to oil 6:1, NaOH 1.0 %wt of oil
<b>Conceptual design</b>	
<i>General requirements:</i>	
Form and operation	simple, constructed and operated conveniently
Glycerol separation during reaction	not applied
<i>Mixing requirements:</i>	
Application	liquid-liquid reaction, low-viscosity blending
Mixing intensity	sufficient but not excess
Homogeneity	homogeneous concentration field
<b>Detailed design</b>	
<i>Design condition:</i>	
Residence time	each unit = 28.5 sec, total = 2.85 min
Predicted purity of ME	96.5 %wt
<i>Assumptions:</i>	
Physical properties of fluid	based on ME
Volumes of turbines, separating plates, baffles, and shaft	neglected during design
<i>Constraints:</i>	
Construction material	3 inch API pipe
Motor speed	capable range of 0-1,000 rpm

in all calculations. The volumes of turbines, baffles, shaft, and separating plates were unknown during the design so that they were neglected in all calculations.

2) **Constraints:** The diameter of the reactor was specified based on an available 3-inch API pipe for convenient construction. Although the stirrer speed at the design flow rate could be estimated, the actual optimum stirrer speed might deviate from the estimated value. Moreover, the experiments were also planned to run at various

flow rates less than the designed flow rate. The optimum stirrer speeds might be lower than the designed stirrer speed. So, the designed stirrer speed was preferred to be in the middle of the capable speed range of an available motor ( $\approx 500$  rpm) to ensure that the optimum stirrer speeds of all experiments would be in the capable speed range of the motor.

3) **Tank and turbine geometries:** As discussed, the required mixing intensity changed along the degree of conversion of reaction. If this



effect was included for design turbine diameters in all units, the optimum diameters might be obtained. However, more information was necessary for the calculations. The problem could be dramatically simplified by neglecting this effect so that all turbines were preferred to have the same diameter. Most of the tank and turbine geometries of each unit were designed based on the typical proportions of a top-entering mixer. The ratio of turbine diameter to tank diameter ( $D_d/D_t$ ) was adjusted to meet the stirrer speed constraint. The location of the turbine was adjusted by placing it at the middle height of each unit. The opening area of the separating plate to allow for the shaft was designed to be as small as possible to prevent the inter-stage mixing.

#### 4) Power supply and stirrer speed:

Although there was no reported information for specifying the power supply for transesterification of RPO, the problem was simplified by specifying the power supply based on a rough quantitative way for low-viscosity blending. McCabe *et al.* (1993) suggested that power supply in the ranges of 0.1 to 0.2 kW/m<sup>3</sup> provides mild mixing, 0.4 to 0.6 kW/m<sup>3</sup> provides vigorous mixing, and 0.8 to 2.0 kW/m<sup>3</sup> provides intense mixing. In this design, the power supply of 2.0 kW/m<sup>3</sup> was chosen.

Another design parameter of interest was stirrer speed. For a given system (fluid properties, geometries, and power supply), the stirrer speed could be calculated through the interrelation of 2 dimensionless numbers characterizing a stirred tank as follows.

$$N_p = \frac{P g_c}{\rho N^3 D_a^5} \quad (2)$$

$$N_{Re} = \frac{N D_a^2 \rho}{\mu} \quad (3)$$

where  $N_p$  is the power number;  $N_{Re}$  is the Reynolds number;  $P$  is the impeller horsepower (Watt);  $g_c$  is the Newton's-law proportionality factor;  $\rho$  is the fluid density (kg/m<sup>3</sup>);  $N$  is the impeller speed (1/s);  $D_a$  is the impeller diameter (m); and  $\mu$  is the fluid dynamic viscosity (Pa.s). The available empirical correlation of  $N_p$  and  $N_{Re}$  for standard

baffled tank fitted with 6-blade disk turbine at  $D_d/D_t$  of 0.33 (McCabe *et al.*, 1993) was extended to roughly estimate the stirrer speeds at various ratios of  $D_d/D_t$ .

5) **Mixing time:** The mixing time was estimated based on a complete mixing model for low to medium viscosity blending in which an essentially complete mixing (99%) could be achieved when the content of a tank was circulated for about five times. The mixing time was calculated from a correlation equation of 6-blade disk turbine as following (McCabe *et al.*, 1993).

$$t_m = 5 \frac{\pi D_t^2 H}{4} \frac{1}{0.92 N D_a^2 D_t} \quad (4)$$

where  $t_m$  is the mixing time (s);  $N$  is the stirrer speed (1/s);  $D_a$  is the turbine diameter (m);  $D_t$  is the tank diameter (m); and  $H$  is the tank height (m).

### 3.3 The designed continuous reactor

For a given condition (residence time, power supply, and tank diameter), the effect of  $D_d/D_t$  on the stirrer speed ( $N$ ), Reynolds number ( $N_{Re}$ ), and ratio of mixing time to residence time ( $t_m/t_r$ ) was considered to specify the appropriate turbine diameter. The calculations were performed at various ratios of  $D_d/D_t$  in the range of 0.2 to 0.8 and the results are shown in Table 2. The stirrer speed varied from 315 to 3,232 rpm in which smaller values of  $D_d/D_t$  required the higher stirrer speeds. The  $N_{Re}$  varied from 2,480 to 3,865. All values of  $N_{Re}$  were quite similar from the viewpoint of flow region in which they were in the transition flow region. The obtained values of  $t_m/t_r$  were quite similar in all cases. The mixing time was very much less than the design residence time, which was required to obtain a homogeneous concentration field. The  $D_d/D_t$  at 0.6 was preferred because it met the speed constraint. The important dimensions of each unit of the 6-stage continuous reactor are shown in Table 3.

## Materials and Methods

### 1. Materials

Commercial grades of RPO, methanol

**Table 2.** The effect of  $D_a/D_t$  on  $N$ ,  $N_{Re}$ , and  $t_m/t_r$  at the given condition (residence time of each unit of 28.5 sec, power supply of 2.0 kW/m<sup>3</sup>, and tank diameter of 8.1 cm).

$D_a/D_t$	$N$ (rpm)	$N_{Re}$	$t_m/t_r$
0.2	3,232	2,480	0.07
0.3	1,633	2,820	0.06
0.4	1,005	3,085	0.06
0.5	693	3,325	0.05
0.6	523	3,520	0.05
0.7	394	3,700	0.05
0.8	315	3,865	0.05

( $D_t$ ) tank diameter, ( $D_a$ ) turbine diameter, ( $N$ ) stirrer speed, ( $N_{Re}$ ) Reynolds number, ( $t_m$ ) mixing time, ( $t_r$ ) residence time.

**Table 3.** The important dimensions of each unit of the designed 6-stage continuous reactor.

Tank diameter	81.0 mm
Tank height	81.0 mm
Impeller diameter	48.0 mm
Blade width	10.0 mm
Blade length	12.0 mm
Baffle width	7.0 mm
Impeller location height	40.5 mm
Opening area of a separating plate (excluding the shaft area)	4% of cross-area

(95%), and NaOH (98%) were used. The free acid content of the oil was determined according to AOCS official method Ca 5a-40 (Link, 1989) to be 0.02 % wt of oil. Pro-analysis grade of hexane, diethyl ether, formic acid, and benzene were used in the analysis.

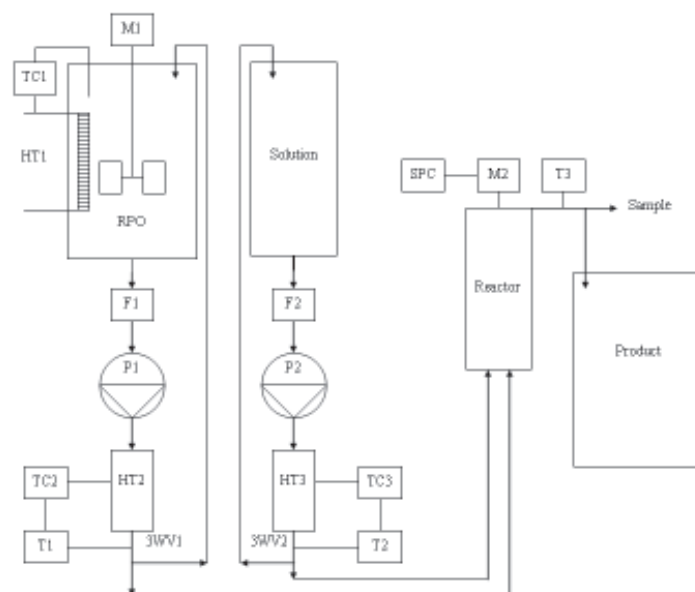
## 2. Experimental equipment

A schematic diagram of the experimental set up is shown in Figure 2. Two storage tanks were used to store RPO and methanol/NaOH solution. Peristaltic pumps model 323 E/D and model 401 U/D (Watson Marlow) were used to pump RPO and methanol/NaOH solution. Two in-line flow meters were used to measure the flow rates of the reactants. A 1 kW heater, a stirrer, and a temperature control unit were installed in the RPO storage tank to preheat RPO. A 3.5 kW in-line heater was

installed in each line of the reactant. Temperature probes connected with temperature control units were installed at the inlet ports of the reactor to measure and control inlet temperatures of the reactants. The heating systems can control the inlet reactant temperatures to within  $\pm 0.4^\circ\text{C}$  of the set point. The total volume of the reactor was measured to be 2.272 liters. The reactor was insulated to prevent heat loss. The product was discharged from the reactor at the top via an overflow discharge port. The samples of product were collected through the sampling port. The remaining product was stored in a product storage tank.

## 3. Experimental procedures

Initially, the empty reactor was pre-heated. The two feed pumps were turned on to circulate



**Figure 2. Schematic diagram for continuous transesterification of RPO. (Mn), motors; (Fn), in-line flow meters; (Pn), peristaltic pumps; (HTn), heaters; (TCn), temperature control units; (Tn), temperature probes; (SPC), speed control unit; (3WVn), 3-way valves.**

reactants from storage tanks through in-line flow meters, in-line heaters, and return to storage tanks. During this time, the pump flow rates were preset. Next, all heaters were turned on and pre-start operation was run until the inlet temperatures of the reactants were constant. After that, reactants were fed into the reactor. When the whole reactor was filled, the stirrer motor was turned on and set at the required speed. Pump flow rates were then finely adjusted. When the desired operation condition was achieved, timing was started. During the experiment, reactant flow rates, reactant inlet temperatures, product temperature, and stirrer speed were monitored. A sample of about 5 ml of the product was collected at 3 times of replacement of the reactor for all experiments. The sample was immediately washed with a large amount of water to stop the reaction. The ester phase was then centrifuged and kept for further analysis.

#### 4. Analysis

The sample was analyzed by thin layer chromatography/flame ionization detector (TLC/

FID). Analysis was performed using an Itronscan MK-6s with Chromarods type S-III quartz rod (Mitsubishi Kagaku Iatron). The sample was diluted in hexane and 1  $\mu$ l of the solution was spotted on the rod. The rods were developed in hexane/diethyl ether/formic acid (50:20:0.3 vol/vol/vol) for 8 cm and in hexane/benzene (1:1 vol/vol) for 10 cm. The rods were dried and scanned under the following conditions: hydrogen flow rate of 160 ml/min, air flow rate of 2.0 l/min, speed of 30 s/scan. Triglycerides, diglycerides, monoglycerides, fatty acid, and ME were effectively separated. Peak areas were calculated with a chromatography data system, ChromStar. Their weight percentages (%wt) on a glycerol-free basis were determined.

#### 5. Power consumption

The net power (excluding friction and inertia power) delivered to the reaction mixture as the mixing intensity was measured by a dynamometer. A force gauge model FG-5000A (Lutron) was used to measure the motor torque and a tachometer



model DT-240P (Digicon) was used to measure the rotational speed. The measurement was repeated 3 times for each condition.

### Results and Discussion

The performance test was run at the designed reaction condition. Residence time was varied at 3, 6, 9, and 12 minutes and stirrer speed was varied in the range of 100-800 rpm at every 100 rpm. The experiment was run at least 2 times for each condition. The important designed and experimental results are summarized in Table 4. The optimum stirrer speeds providing the highest purities of ME for all flow rates were in the speed range of the motor and lower flow rates required lower stirrer speeds. A comparison between the experimental results at the residence time of 3 min and the designed results at the residence time of 2.85 min showed that the optimum stirrer speed was very close to the designed result. The power consumption was around 1.4 times of the estimated power. Interestingly, the optimum ratios of  $t_m/t_r$  for all flow rates were similar and they were very close to the estimated value. The power consumption of a stirrer in the range of 0.2-0.6 kW/m<sup>3</sup> was sufficient for producing saleable biodiesel and the required power is much increased when the reactor was operated at the shorter residence time.

The RTD test was performed twice at the shortest residence time that could produce saleable

biodiesel (6 min) and at the optimum stirrer speed of 300 rpm. The negative step change method (Paul *et al.*, 2004) was used. The heated RPO (60°C) was used as the main fluid and a small amount of ME was used as a tracer. The flow rate of RPO was set to obtain the desired residence time. Initially, RPO and ME were fed into the reactor for an extended period to obtain a constant tracer concentration in an outlet stream ( $C_{out} = C_{in} = C_0$  for time  $t < 0$ ). At time  $t = 0$ , the tracer supply was stopped so that  $C_{in} = 0$  for time  $t \geq 0$ . The samples of the outlet stream were collected at pre-specified time intervals and analyzed by TLC/FID to determine the weight concentration of the tracer ( $C_t$ ). A washout function  $W(t)$ , normalized tracer concentrating is defined as shown in equation 5. A dimensionless variance ( $\sigma^2$ ) and the number of ideal CSTR in series ( $n$ ) were calculated by equations 6 and 7. The experimental RTD data are shown in Figure 3. The calculated  $n$  was 5.98, which it was very close to the 6-ideal CSTRs in series as expected and required.

$$W(t) = C_t / C_0 \quad (5)$$

$$\sigma^2 = \frac{2 \int_0^{\infty} tW(t)dt}{\left[ \int_0^{\infty} W(t)dt \right]^2} - 1 \quad (6)$$

$$n = \frac{1}{\sigma^2} \quad (7)$$

**Table 4. The important designed and experimental results for transesterification of RPO in a 6-stage continuous reactor at molar ratio of methanol to oil of 6:1, temperature of 60°C, and NaOH of 1.0 %wt of oil.**

Parameter	Total Residence Time (min)				
	2.85 <sup>a</sup>	3 <sup>b</sup>	6 <sup>b</sup>	9 <sup>b</sup>	12 <sup>b</sup>
Purity of ME (%wt)	96.5	95.8	97.5	98.2	99.2
Optimum stirrer speed (rpm)	523	500	300	200	200
$N_{Re}$ (based on ME)	3,520	3,370	2,020	1,350	1,350
Mixing time/Residence time	0.05	0.04	0.04	0.04	0.03
Power supply per unit volume (kW/m <sup>3</sup> )	2.0	2.8	0.6	0.2	0.2

<sup>a</sup>Designed result, <sup>b</sup>Experimental result

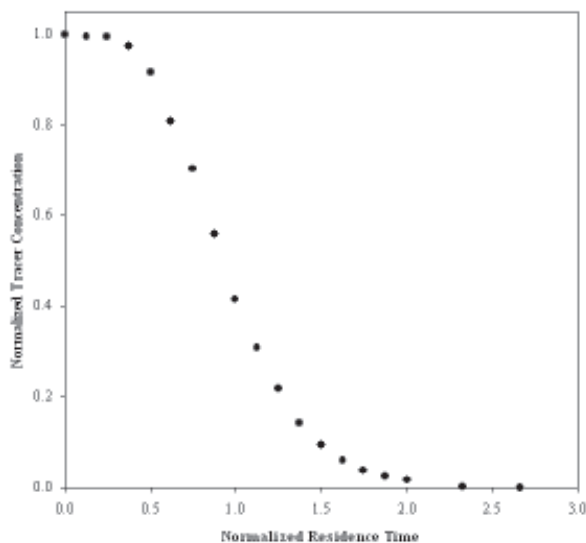


Figure 3. The RTD of a 6-stage continuous reactor at residence time of 6 min and stirrer speed of 300 rpm.

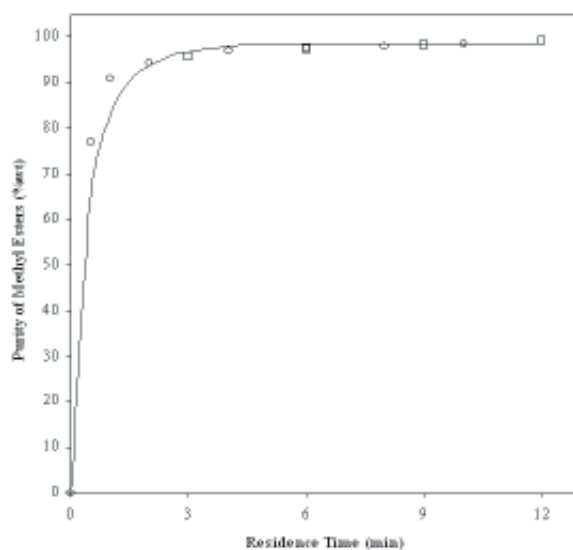


Figure 4. The production performances for transesterification of RPO at molar ratio of methanol to oil of 6:1, temperature of 60°C, and NaOH of 1.0 %wt of oil. (○) experimental result of a well-mixed batch reactor (Leevijit *et al.*, 2004a); (line) predicted result of a 6-stage continuous reactor (Leevijit *et al.*, 2004b), (□) experimental result of a 6-stage continuous reactor.

The production performance of the continuous reactor was plotted and compared to the batch reaction from our previous work (Leevijit *et al.*, 2004a) as shown in Figure 4. Both results are

very much in agreement. At same reaction times, same purities of ME were produced. This result revealed that, in the studied range, the production performance of the continuous reactor was equi-

valent to a PFR as required. Figure 4 also shows the comparison between the predicted production performance (Leevijit *et al.*, 2004b) and the experimental production performance of the continuous reactor. The reaction modeling could accurately predict its production performance. Vassilatos and Toor (1965) and McKelvey *et al.* (1975) also found that the reaction modeling based on a homogeneous concentration field with reaction kinetics could provide accurate prediction for a slow reaction relative to the mixing rate. For this class of reaction, there is plenty of time to complete the mixing before the reaction makes any significant progress. As a consequent, in the studied range, it could be concluded that the designed process mixer could supply sufficient mixing relative to reaction rate and flow rate in which a homogeneous concentration field was obtained as required.

The production yield neglecting the losses due to glycerol separation and washing processes was measured 3 times at the residence time of 6 min. The product of the reaction mixture was collected and the collecting time was measured. The product was suddenly washed by hot water and further washed until it was clean in which glycerol, methanol, NaOH, and other impurities were completely removed from the product. The product was heated to 120°C to remove water and cooled down to room temperature. The weight of the final product was measured. The production yield of the final product based on a mass flow rate of the fed RPO was calculated, and the production yield of produced ME based on a mass flow rate of the fed RPO was also calculated. The average production yield by weight of the final product to the fed RPO was 94.7% and the average production yield by weight of the produced ME to the fed RPO was 92.3%.

### Conclusion

An effective continuous reactor for transesterification of RPO with methanol was designed in the form of a 6-stage mechanically stirred tank. Rushton turbines, 4 standard baffles, and plates

with a small opening were installed inside. The benefit of this design was that it had a simple form, which could be conveniently constructed and operated.

The performance test revealed that the reactor could work well without problems. At the design reaction condition, the reactor could produce ME at purities of 97.5-99.2 %wt within residence times of 6-12 min. Its production performance was found to be equivalent to a PFR. The power consumption of a stirrer in the range of 0.2-0.6 kW/m<sup>3</sup> was required. Moreover, the reaction modeling based on homogeneous concentration field and reaction kinetics could accurately predict its production performance. The production yield by weight of the final product to the fed RPO was 94.7% and the production yield by weight of the produced ME to the fed RPO was 92.3%. The developed continuous reactor has good potential for producing ME to be used as biodiesel.

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