

A NOVEL CONTINUOUS-FLOW REACTOR USING REACTIVE DISTILLATION FOR BIODIESEL PRODUCTION

B. B. He, A. P. Singh, J. C. Thompson

ABSTRACT. *The production of biodiesel through batch and existing continuous-flow processes requires the use of a much higher excess alcohol, typically 100%, than the stoichiometric molar requirement in order to drive the transesterification reaction to completion. This excess alcohol must be recovered in a separate process, which involves additional capital and operating costs. In this study, a novel reactor system using reactive distillation (RD) was developed and investigated for biodiesel preparation from canola oil and methanol. The goal was to significantly reduce the use of excess methanol while maintaining a high methanol:glyceride molar ratio inside the RD reactor by recycling a small amount of methanol within the system. Reactant conversion rate and product yield were used as the criteria for the reactor evaluation. The effect of the methanol:glyceride ratio was studied on a laboratory-scale perforated-tray RD reactor system. Product parameters such as methyl ester content, glycerides, and methanol content were analyzed. Preliminary results showed that the RD reactor with a methanol:glyceride ratio of 4:1 (molar), in which the use of methanol was cut down by 66%, gave a satisfactory biodiesel yield and oil conversion rate at a column temperature of 65 °C. Total reaction time in the pre-reactor and RD column was about 3 min, which is 20 to 30 times shorter than in typical batch processes. The productivity of the RD reactor system was about 6.6 m³ biodiesel per m³ reactor volume per hour, which is 6 to 10 times higher than that of batch and existing continuous-flow processes.*

Keywords. *Biodiesel, Canola oil, Fatty acid alkyl esters, Reactive distillation, Transesterification.*

Biodiesel from vegetable oils has attracted increasing interest among government, industry, and the general public. It has been shown to be the best supplement to fossil-based fuels due to environmental advantages, renewable resource availability, and the ability to lessen the dependence on imported oil. Currently, the dedicated production capacity is estimated to be between 60 and 80 million gallons per year, and it has been reported that up to 200 million gallons of production capacity are available through long-term production agreements with existing biodiesel marketing firms (National Biodiesel Board, 2004). As more and more commercial biodiesel production facilities come under consideration, the demand for a technically and economically sound reactor technology increases.

Biodiesel can be made from transesterification of vegetable oils or animal fats with methanol or ethanol in the presence of a catalyst. Studies of the mechanism and kinetics have shown that this process consists of a number of consecutive, reversible reactions (Noureddini and Zhu, 1997; Darnoko and Cheryan, 2000a; Freedman et al., 1986). Triglycerides are first reduced to diglycerides, and then to monoglycerides. Lastly, the monoglycerides are reduced to

fatty acid esters and glycerol, as shown in figure 1, where the TG, DG, MG, and GL stand for tri-, di-, monoglycerides, and glycerol, respectively; R₁, R₂, and R₃ are the same or different alkyl groups of typically C₁₄-C₂₂ chains; and R₄ is the methyl or ethyl group depending on the alcohol used. The completion of the reaction depends on multiple parameters, including the methanol:glycerol molar ratio, catalysts, temperature, reaction time, and properties of the feedstock. As an effective means of conversion enhancement, excess alcohol is used to drive the equilibrium towards the product side. In the biodiesel industry, the molar ratio of alcohol to glyceride is typically 6:1, or twice as much as the stoichiometric molar ratio of 3:1.

The excess alcohol remaining in the esters and glycerol after reaction must be recovered and purified for reuse. Batch processes are labor and energy intensive and low in productivity. Continuous transesterification processes are preferred over batch processes in large-capacity commercial production due to the consideration of consistent product quality and low costs of capital and operation per unit of product. Continuous transesterification of vegetable oils to mono-alkyl esters was proposed as early as in the 1940s (Trent, 1945; Allen et al., 1945) and studied until recent years (Noureddini et al., 1998; Darnoko and Cheryan, 2000b; Antolin et al., 2002; Peterson et al., 2002; Van Gerpen and Knothe, 2005). Most of the existing processes still utilize 100% (by molar quantity) or more excess alcohol. The operating parameters, such as reaction time of 60 to 120 min and operating temperatures of 20 °C to 70 °C, were directly adopted from batch operation processes.

Application of the reactive distillation technique to biodiesel production can lead to a more effective and efficient biodiesel reactor system. Reactive distillation (RD)

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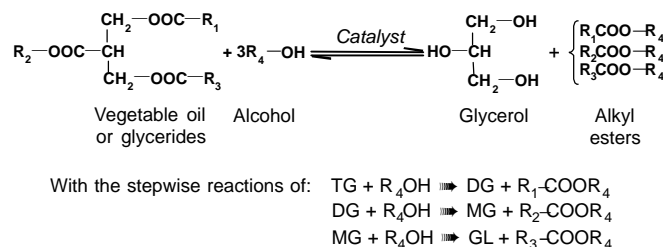


Figure 1. Transesterification of vegetable oils to produce fatty acid alkyl esters.

is a chemical unit operation in which chemical reactions and product separations occur simultaneously in one unit. It is an effective alternative to the classical combination of reactor and separation units, especially when reversible or consecutive chemical reactions are involved. Both packed and tray columns can be used for RD applications; however, tray columns are recommended for homogenous systems because of the greater liquid hold-up and the relatively longer retention time. Application of reactive distillation in chemical engineering is not a new practice. Extensive research has been conducted on its applications in the chemical industry, and it is increasingly being used in various chemical systems (e.g., Solokhin and Blagov, 1996; Malone and Doherty, 2000; Tuchlenski et al., 2001). In some applications, particularly when reversible reaction equilibrium prevents high conversions, the RD technique can be employed to remove the reaction products from the reaction zone to improve overall conversion rate and selectivity. High chemical conversion rates and low capital and operating costs are among the significant advantages of RD over the conventional sequential process. Examples of RD application include esterification, ester hydrolysis, and ether synthesis (Mohl et al., 1999; Polpken et al., 2001; Omota et al., 2003).

The ultimate goal of this research project was to explore a technically and economically sound reactor technology for commercial biodiesel production by utilizing the RD technique. The specific goal of this study was to test the applicability of a laboratory-scale continuous-flow RD reactor for biodiesel preparation. The objectives were: (1) to prepare fatty acid esters by employing the RD technique, (2) to reduce the use of excess alcohol in the feed but create a high methanol:glyceride ratio in the reaction zone by recycling a small quantity of alcohol, and (3) to greatly increase the biodiesel productivity per unit reactor volume per unit time through the combination of elevated operating temperature (65°C in the column) and much-shortened reaction time.

MATERIALS AND METHODS

CHEMICALS AND REAGENTS

Crude canola oil and methanol were used as the feedstock in this research. Potassium hydroxide was used as the catalyst. The canola oil was produced at the oil seed processing plant at the Department of Biological and Agricultural Engineering of the University of Idaho using well-established procedures (Peterson et al., 1983). The oil was screw-pressed and filtered to remove any particulates. No further treatments were performed on the oil. The acid value of the oil was 1.97 mg KOH/g sample. The fatty acid profile of the canola oil was analyzed using the GC method

Table 1. Fatty acid profile of canola oil used in this research.

Fatty Acids	Composition (% wt)		
	Test 1	Test 2	Average
Palmitic (16:0)	4.4	4.5	4.5
Stearic (18:0)	1.8	1.8	1.8
Oleic (18:1)	60.9	60.5	60.7
Linoleic (18:2)	19.1	19.1	19.1
Linolenic (18:3)	9.5	9.5	9.5
Eicosic (20:1)	1.8	1.8	1.8
Erucic (22:1)	0.8	1.0	0.9

(Hammond, 1991) and is listed in table 1. Methanol (analytical grade) and potassium hydroxide (ACS certified, purity >87.9% wt), were from J.T. Baker (Phillipsburg, N.J.). Other analytical reagents and standard chemicals, i.e., triolein, diolein, methyl oleate, and glycerol, were all analytical grade and purchased from Sigma-Aldrich Co. (St. Louis, Mo.).

EQUIPMENT AND EXPERIMENTAL SETUP

A laboratory-scale continuous-flow RD reactor system (fig. 2) was developed and tested. The central component was a glass distilling column with perforated Oldershaw plates (ChemGlass, Vineland, N.J.). This 10-plate column had an inner diameter of 28 mm, a weir height of 1.7 mm, and a distance of 25 mm between plates. The nominal liquid hold-up on each plate was 1 mL, totaling 10 mL for the column. The column had a vacuum jacket around it as a means of thermal insulation to prevent heat loss from the column during operations.

A short 150 mm (5 mL capacity) in-line static mixer (Cole-Parmer, Vernon Hills, Ill.) was used as a feed mixer, heat exchanger, and pre-reactor prior to the RD column. The lower end of the column was fitted to a 500 mL three-neck round-bottom flask as the reboiler. A water-cooled condenser was fitted to the top of the column to recover alcohol. The feed entered the column near the top. The condensed methanol was combined with the feed methanol and refluxed

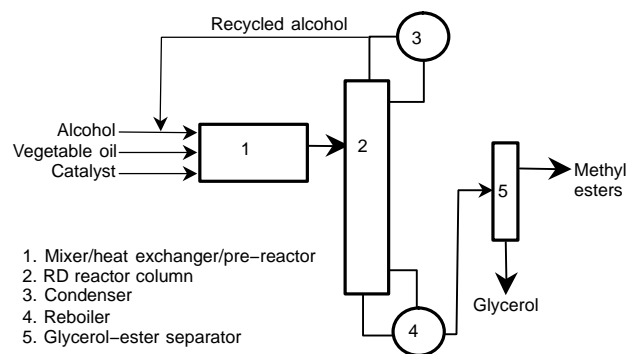


Figure 2. Schematic of RD reactor system.

back to the column. The product mixture was withdrawn from the reboiler to a glycerol-ester separator, where the glycerol and esters were separated by gravity in a continuous mode. The separator was 70 mm in diameter \times 300 mm long with an adjustable entry point. The input and output streams of methanol/KOH solution, canola oil, and product mixture were handled simultaneously with three Masterflex peristaltic pumps (Cole-Parmer, Vernon Hills, Ill.), which were calibrated and adjusted to achieve the desired flow rates ranging from 0.5 to 8.0 mL/min. Temperatures of the pre-reactor, condenser, column, and reboiler were measured and monitored. The heating rate of the reboiler was controlled as a constant in order to maintain a constant methanol recycling rate. The heating load and feedback control were accomplished with a Fuji PXR3 and PXR4 PID/feedback controller (distributed by TTI, Inc., Tokyo, Japan).

PRODUCT CHARACTERIZATION

The raw product sample taken directly from the reboiler was treated with 0.1 N HCl to stop the reaction and was analyzed for methyl ester content and glycerides without further treatment. The methanol content in the product was determined from the product mixture accumulated in the separator at the end of the operation.

Compositions of the methyl esters and mono-, di-, and triglycerides in the samples were analyzed by an HP 1090 HPLC with Altech 2000 ELSD (Alltech Associates, Inc., Deerfield, Ill.). The column, C18 with 7 μ m SGX, was maintained at 40 °C, while the ELSD was maintained at 60 °C. The flow rate of nitrogen, the carrying gas, was set at 1.5 L/min. Gradient elution was conducted with aqueous-organic and non-aqueous mobile phase steps: 70% acetonitrile and 30% water at 0 min, 100% acetonitrile at 10 min, and 50% acetonitrile and 50% 2-propanol-hexane (5:4 v/v) at 20 min with 7.5 min hold-up time (Komers et al., 2001).

The methanol content in the biodiesel samples was determined by the difference of a sample before and after heating it to 90 °C for 30 min to drive off the residual alcohol.

TERMS USED IN THIS STUDY

To avoid ambiguity, the **reactor volume** in this study includes the liquid volume held in the pre-reactor (5 mL) and on the plates of the RD column (total 10 mL). The liquid volume held in the reboiler was about 100 mL (the relative amount of liquid in the reboiler vs. that in the column could be considerably reduced in larger-scale setups). The **reaction time** in this study, therefore, was calculated as the liquid retention time in the pre-reactor and column. **Productivity** refers to the amount of fatty acid esters (m³) produced per unit reactor volume (m³) per unit time (h). Two major process parameters used to examine the efficiency of the RD system were the conversion rate and yield. **Conversion rate** refers to the percentage of oil glycerides (the reactant) converted to the intermediate and end products versus the total amount in the feed, namely:

conversion rate (%)

$$= \frac{\text{glycerides converted (g or mol)}}{\text{glycerides in the feed (g or mol)}} \times 100\%$$

$$= \left(1 - \frac{\text{glycerides left after reaction (g or mol)}}{\text{glycerides in the feed (g or mol)}} \right) \times 100\%$$

Yield is defined as the actual amount of target product (fatty acid esters or biodiesel) formed, out of the maximum (theoretical) amount possible, from the oil glycerides, expressed as a percentage:

$$\text{yield (\%)} = \frac{\text{Actual FA esters formed (g or mol)}}{\text{Maximum amount possible (g or mol)}} \times 100\%$$

RESULTS AND DISCUSSIONS

OPERATION OF THE RD REACTOR

Generally speaking, the operation of an RD reactor is complicated because its performance is affected by several parameters, including the reaction kinetics, size of the reaction and separation zones, reflux ratio, feed rate, and feeding tray location, etc. (Solokhin and Blagov, 1996; Tuchlenski et al., 2001). The optimum operating conditions are determined as the result of systematic investigations of all operating parameters. However, the complexity of the RD reactor system is minimized when applied to the biodiesel production for four reasons. First, the difference between the boiling temperatures of methanol and fatty acid esters is so large that the separation of the excess alcohol from the product mixture becomes easy (table 2). The rectification zone of the RD column can be easily determined, which requires one or two theoretical separating stages. Second, as a result of the low rectification requirement, the feeding position is then determined to be at the upper section of the RD column, one or two theoretical separating stages below the top plate. Third, the transesterification reaction occurs in the liquid phase only. The plates below the feeding point determine the reaction zone. The reaction time is then established by the total liquid hold-up and the feeding rate of the reactants. Finally, the operating temperature is easily determined by the boiling temperature of the alcohol used when the operating pressure of the column has been determined.

According to the above analysis on the application of RD to the transesterification of canola oil and methanol, a laboratory-scale RD reactor system was established, as shown in figure 2. Prior to each experiment, the catalyst, potassium hydroxide (KOH), was dissolved in methanol to

Table 2. Boiling temperatures of chemical components in the transesterification of vegetable oils to biodiesel.

Component	Boiling Point (°C)
Methanol	64.7 ^[a]
Glycerol	290.0 ^[a]
Methyl palmitate	196 (at 15 torr) ^[a]
Methyl stearate	215 (at 15 torr) ^[a]
Methyl oleate	190 (at 10 torr) ^[a]
Methyl linoleate	215 (at 20 torr) ^[b]
Methyl erucate	221 (at 5 torr) ^[b]
Ethyl palmitate	227 (at 757 torr) ^[a]
Ethyl stearate	201 (at 10 torr) ^[a]
Ethyl oleate	216 (at 15 torr) ^[a]
Methyl esters, rapeseed	369.0 ^[c]
Methyl esters, canola	338.1 ^[c]

^[a] Perry and Green, 1997.

^[b] Weast et al., 1986.

^[c] Goodrum, 2002.

such a concentration that once mixed, the catalyst was 1% wt of the canola oil in the feed. The concentration was determined by the methanol:glyceride ratio used in the experiment. The canola oil and methanol/catalyst solution were pumped quantitatively at a suitable flow rate to avoid flooding or weeping in the RD column. Recycled methanol from the condenser was combined with the reactants before entering the in-line static mixer that also served as a heat exchanger. Heating tape wrapped around the mixer was used to achieve the desired feed oil temperature, i.e., 50 °C in this study. By nature, the mixer/heat exchanger also functions as a reactor in the presence of the reactants and a catalyst. For reactions that are severely equilibrium-limited, such as in the transesterification of canola oil and methanol, the initial reaction rate is high. The reaction rate declines abruptly in a short period of time when the composition of reactants and products approaches equilibrium. In such cases, a reactor upstream of the RD column can handle a substantial part of the reaction, which greatly enhances the RD process. As determined in a separate study (data not shown), the oil conversion rate in the pre-reactor ranged from 30% to 60%, depending on the operating conditions (He et al., 2006). Therefore, the unit serves a triple function as a reactant mixer, heat exchanger, and pre-reactor.

The reactant mixture entered the RD column at the top, flowed across the plate, and then flowed downward through the downcomer to the next plate. While flowing down to the reboiler in a zig-zag manner, the liquid was held on top of each plate by the upward-flowing methanol vapor through the perforated holes. The upward-flowing methanol vapor also served as an agitator in the reactant mixture, providing uniform mixing while it was bubbling through the liquid phase on each plate. As a result, the reaction zone of the RD column consisted virtually of a series of mini-reactors (fig. 3).

Once the product mixture reached the reboiler, operated at 95 °C to 150 °C in this study, most of the unreacted methanol (approx. 90% or more) was vaporized off the product mixture and flowed up through the column constantly. Upon reaching the condenser at the top of the RD column, the methanol vapor was condensed and combined with the feed methanol before re-entering the column. It was this portion of the recycled alcohol that created a high local methanol:glyceride ratio and drove the transesterification reaction in the mini-reactors toward completion. However, this ratio was not directly monitored through composition

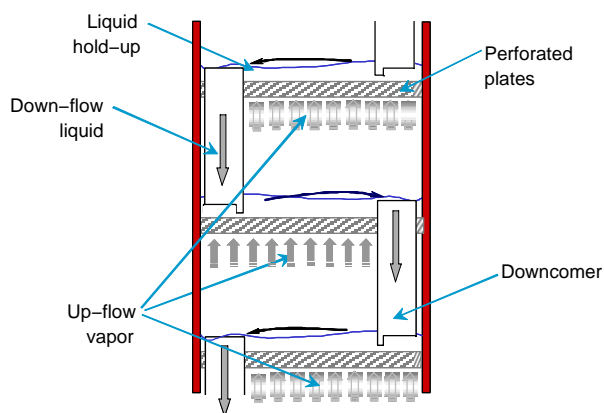


Figure 3. Illustration of the liquid and vapor flows in the RD column.

measurement due to the limitation of the RD reactor construction (silicon glass column). As estimated indirectly through measurement of the methanol condensation rate (data not shown), the overall methanol:glyceride ratio at the feeding plate (the first mini-reactor), after considering a 30% to 60% reaction in the pre-reactor, could be as high as 15:1 to 20:1. As a result, the use of excess alcohol in the feed was considerably reduced.

Product was continuously withdrawn from the reboiler through a pump to a glycerol-ester separator, where the crude glycerol was separated by gravity from the methyl esters. The pumping was controlled in such a way that the working volume was kept constant during steady-state operation.

The RD reactor system was relatively easy to operate compared with its applications in other systems (Solokhin and Blagov, 1996; Malone and Doherty, 2000; Tuchlenski et al., 2001). Each experiment lasted 300 to 360 min, and reactor operation typically required about 60 min to reach steady state before any sampling or observations could be taken.

EXPERIMENTAL RESULTS

Two sets of preliminary trials with and without a pre-reactor were conducted with the RD setup (fig. 2). The main process parameters examined in this study were: the methanol:glyceride ratio, the feed flow rates, and the reaction temperatures. In each trial, stock alcoholic KOH and canola oil were prepared and stored in separate holding reservoirs at room temperature. The methanol:glyceride molar ratios tested were 3.0:1, 3.5:1, 4.0:1, and 4.5:1, which were achieved by adjusting the alcohol/catalyst flow rate relative to that of the canola oil. The column temperature was maintained at 65 °C, which was close to the boiling point of methanol. The operating parameters of the experiments are summarized in table 3. The product composition of total methyl esters, glycerides, and methanol content under different experimental conditions are summarized in table 4.

As expected, the system with the pre-reactor gave higher methyl ester content compared to those without a pre-reactor. The increase in methyl ester content with the increases of methanol:glyceride molar ratio justifies the need for excess alcohol to drive the reaction to a higher yield. The methyl ester content in the product increased from 75.6% to 91.5% as the molar ratio increased from 3:1 to 4.5:1. This 16% relative increase strongly suggests that a higher methanol:glyceride ratio in the reaction zone is preferable. The methyl ester content of the product from the runs without the pre-reactor showed a similar trend. The results also showed

Table 3. Process parameters used in the preliminary experiments.

Trial	Feed Rate (mL/min)			Temperature (°C)		
	MeOH: Glyceride	Canola Oil	MeOH/ KOH	Pre-reactor	Column	Reboiler
With pre-reactor						
1	3.0:1	6.67	0.8	50	65	150
2	3.5:1	5.67	0.8	50	65	125
3	4.0:1	4.95	0.8	50	65	115
4	4.5:1	4.45	0.8	50	65	110
Without pre-reactor						
5	3.0:1	6.67	0.8	50	65	135
6	3.5:1	5.71	0.8	50	65	126
7	4.0:1	5.00	0.8	50	65	123
8	4.5:1	4.45	0.8	50	65	95

Table 4. Summary of preliminary experimental results.^[a]

Trial	Composition (% wt)				MeOH (% wt)	Conversion (%)	Yield (%)
	ME	MG	DG	TG			
With pre-reactor							
1	75.6 ±0.88	14.0 ±0.81	6.85 ±0.59	3.53 ±0.51	2.1	78.4 ±0.71	75.3 ±0.87
2	77.3 ±0.61	3.14 ±0.62	7.44 ±0.60	12.1 ±0.14	1.4	78.2 ±0.54	77.0 ±0.60
3	94.8 ±0.08	1.15 ±0.01	2.04 ±0.09	2.00 ±0.04	0.9	95.1 ±0.08	94.4 ±0.08
4	91.5 ±0.05	2.27 ±0.04	3.34 ±0.05	2.88 ±0.05	1.5	92.1 ±0.04	91.1 ±0.05
Without pre-reactor							
5	62.2 ±2.00	10.4 ±0.95	9.04 ±0.44	18.3 ±0.61	3.7	64.5 ±1.78	61.9 ±1.96
6	77.5 ±0.16	3.64 ±0.16	5.46 ±0.09	13.4 ±0.09	2.1	78.4 ±0.14	77.2 ±0.16
7	93.5 ±0.51	1.52 ±0.12	2.21 ±0.20	2.81 ±0.21	2.0	93.8 ±0.48	93.0 ±0.51
8	94.7 ±0.94	2.24 ±0.95	1.14 ±0.07	1.94 ±0.02	4.9	95.1 ±0.78	94.3 ±0.94

^[a] ME = content of canola fatty methyl esters in product, MG = total monoglycerides in product, DG = total diglycerides in product, TG = total triglycerides in product, MeOH = methanol in product, Conversion = feed oil conversion rate, and Yield = methyl ester yield on feed oil. The data are in "average ±standard deviation" format, except for methanol content in which one mixed sample was measured to give a representative value.

that there was a difference of 14% in conversion rates and yields between the two setups at a low methanol:glyceride ratio, which proved the advantage of using a pre-reactor. But the difference was not significant once the ratio increased. In summary, use of such a mixer/pre-reactor can increase the conversion rate or reduce the required size of the reaction zone, which potentially reduces the capital and operating costs.

About 5% to 10% of unreacted methanol was pumped out with the product mixture. After the ester and glycerol separation in the separator, about 30% of it ended up in the ester layer, while the remainder was found in the glycerol layer. At a 4:1 molar ratio, methanol content in the biodiesel was found to be the lowest for the conditions tested. The methanol content in the esters increased if the methanol:glyceride ratio was increased to 4.5:1 due to a greater excess of methanol and its incomplete vaporization from the reboiler at the same reboiler heating rate, as indicated by the lowered reboiler temperatures.

Both product yield and feedstock conversion rate were affected significantly by the operating parameters, especially the feed methanol:glyceride ratio. When low methanol:glyceride ratios were used, e.g., 3.0:1 and 3.5:1 (molar), the conversion rates and yields were both low, virtually at similar levels as those in batch reactions with a theoretical methanol:glyceride ratio (Singh et al., 2006). The conversion rate increased to 95.1% and the yield increased to 94.4% if the methanol:glyceride ratio was increased to 4.0:1 when the pre-reactor was used. When the methanol:glyceride ratio was further increased to 4.5:1, the conversion rate and yield were lowered slightly. The results showed similar trends in conversion rates and yields if no pre-reactor was used. Observations showed that there was sufficient methanol being circulated in the RD column with a methanol:glyceride ratio of 4:1. Further increasing the ratio to 4.5:1 would not benefit the conversion rate and yield significantly.

In batch and existing continuous-flow processes, the liquid retention time affects the conversion rate considerably. Long retention time must be used. Quite the opposite is true for the RD system. It was determined, through several trials, that a flow rate of 5 mL/min gave no significant operational difficulties. The liquid hold-up of the RD reactor column was about 10 mL, and the pre-reactor had a volume of 5 mL. The liquid retention time was about 2 min in the RD reactor and about 1 min in the pre-reactor. Therefore, the overall liquid retention time in the reactor system was about 3 min. The

short reaction time led the RD process to a high productivity. Compared to a batch or other continuous-flow reactors, where the retention time was 60 to 90 min (Noureddini et al., 1998; Darnoko and Cheryan, 2000b; Peterson et al., 2002; Van Gerpen and Knothe, 2005), the reaction time was significantly reduced by 20 to 30 times under the current operating conditions. The productivity of the product mixture (5 mL per 15 mL total reactor volume per 3 min) translates to approximately 0.11 mL of product mixture per unit reactor volume per min, or 6.6 m³/m³·h. This productivity is about 6 to 10 times higher than that of batch or existing continuous-flow processes. In other words, to achieve a specified production, the volume or capital cost of the RD reactor system would be greatly reduced. The experimental results have also shown that a molar methanol:glyceride ratio of 4.0:1 was feasible for high feedstock conversion rates and product yields. The use of excess alcohol was cut by 66%, implying that the downstream alcohol recovery effort and/or the operating cost would be reduced by two-thirds as well. In brief, the RD reactor system possesses enormous advantages over batch and traditional continuous-flow reactors in biodiesel production, especially for large-quantity commercial facilities, due to its high productivity and reduced alcohol requirement. However, further systematic investigations are necessary for all operating parameters and their effects on product quality. In our future research, effort will also be placed on improving the RD reactor efficiency and on estimating the energy utilization before the advantages of the RD reactor system can be fully established.

CONCLUSIONS

A novel continuous-flow reactor using reactive distillation, as described in this article, has been found feasible for biodiesel production from canola oil. The operating parameters of 65 °C column temperature and 4:1 methanol:glyceride molar ratio with a pre-reactor have yielded promising results. Preliminary results showed that the RD reactor was very effective in transesterifying canola oil to biodiesel. The use of excess alcohol in the feed was reduced by 66%. This implies that the downstream alcohol recovery effort would also be reduced by 66%. The short reaction time, which was 10 to 15 times shorter than those used in batch and existing continuous-flow reactors, led to a 6 to 10 times higher productivity. In summary, this RD reactor bears three major

advantages over batch and traditional continuous-flow processes: (1) shorter reaction time and higher unit productivity, which is highly desirable in commercial production units; (2) much lower excess alcohol requirement, which greatly reduces the effort of downstream alcohol recovery and operating costs; and (3) lower capital costs due to its smaller size and the reduced need for alcohol recovery equipment. Further investigations are necessary on the process efficiency, energy utilization, and economic analysis in order to fully establish its impact on large-scale commercial biodiesel production. Economic analysis of the RD reactor system will be performed in a scaled-up facility. The overall process integration will also be included to produce a meaningful economic analysis.

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