

# PROCESS MODEL FOR BIODIESEL PRODUCTION FROM VARIOUS FEEDSTOCKS

D. Tapasvi, D. Wiesenborn, C. Gustafson

**ABSTRACT.** Tools are needed to evaluate and compare different available feedstocks, and process parameters and modifications for biodiesel production. To address this need, a biodiesel process model was developed with commonly used spreadsheet software and process-engineering principles. The basis of the model is a continuous process with two stirred-tank reactors and sodium methoxide catalysis. The process was modeled as 27 units with 51 flows and 18 components. Mass flow rates and compositions of the process input and output streams were quantified using mass and component balances, energy balances, stoichiometric relations, and established process parameters. Oil composition and rate, methanol:triglyceride ratio, and expected transesterification of triglyceride were the user-specified inputs in the model. Based on commonly reported parameters (6:1 methanol:triglyceride ratio and 98% transesterification) and a basis of 100 kg/h crude soybean oil, the model computes (in kg/h) inputs of 13.8, 10.8, 34.7 for methanol, 10% sodium methoxide in methanol, and process water, respectively, and outputs of 93.5, 10.3, 55.6 for soy biodiesel, glycerol, and waste stream, respectively. These mass flow rates can be linked to cost data for calculating the material costs from various feedstocks, and can be readily adapted to factor in alternative parameters and units.

**Keywords.** Biodiesel, Ester washing, Mass balance, Process modeling, Spreadsheet, Transesterification, Vegetable oil.

Large harvests of traditional crops, low farm prices, dependence on foreign energy sources, and environmental problems due to combustion of fossil fuels have increased interest in renewable energy sources such as biodiesel. Biodiesel is a diesel fuel replacement produced from domestic renewable resources such as vegetable oils, animal fats, and recyclable cooking oils. Chemically, it is defined as the mono-alkyl esters of long-chain fatty acids derived from lipid sources. The demand for biodiesel is on the increase as more people are becoming aware of its potential to address the above issues. More biodiesel production plants are needed to meet the anticipated demand. The foremost step for setting up a biodiesel production plant is evaluation of the economic feasibility of biodiesel production. These economic feasibility studies must be based on authentic and reliable production models.

Various process engineering principles, such as mass and energy balances, can be utilized to collect and organize the production data by analyzing the biodiesel production process in terms of the various inputs, outputs, and their compositions at each process step. Such information is, in turn, needed for quantifying utility requirements and for

sizing equipment. Thus, performing the mass and energy balance analysis of the full process is a key step towards obtaining economic cost data for performing the economic feasibility studies.

Efforts have been reported to quantify the biodiesel production process (Sheehan et al., 1998), but no detailed, transparent approach has been described to quantify the inputs and outputs in the process. Many economic feasibility study groups have applied the so-called "black box" approach by providing only the overall inputs and outputs of biodiesel production collected from private consultancy firms (Frazier Barnes & Associates, 2003; Van Wechel et al., 2002). Such approaches leave the reader with little understanding of the underlying calculations, and no basis for determining whether advances in process technology were taken into consideration. In order to resolve the above issues and in an effort to provide a complete understanding of biodiesel production, a biodiesel process model is needed.

Such a process-modeling approach has been utilized extensively by the chemical and food process industries for detailed design and feasibility evaluation of production plants. One excellent example of such a process model was shown by Fryer et al. (1997), who modeled a cheese manufacturing plant using spreadsheets. Spreadsheets were chosen because of their user-friendly interface and widespread use. A similar approach would be useful for biodiesel production process.

Therefore, the objectives of this study were to collect detailed biodiesel process information and to develop a biodiesel process model based on that information. The model was to quantify all the mass flows involved in the biodiesel production process by using the mass balance principles and with the help of computer spreadsheets and other software tools. The model also incorporates energy balance principles where necessary to complete the associat-

---

Article was submitted for review in December 2004; approved for publication by the Food & Process Engineering Institute Division of ASABE in September 2005.

The authors are **Dhruv Tapasvi**, Graduate Research Assistant, Department of Agricultural and Biosystems Engineering and Department of Agribusiness and Applied Economics, **Dennis Wiesenborn**, Professor, Department of Agricultural and Biosystems Engineering, and **Cole Gustafson**, Professor, Department of Agribusiness and Applied Economics, North Dakota State University, Fargo North Dakota. **Corresponding author:** Dennis Wiesenborn, Department of Agricultural and Biosystems Engineering, North Dakota State University, 1221 Albrecht Blvd, Fargo, ND 58105-5626; phone: 701-231-7277; fax: 701-231-1008; e-mail: d.wiesenborn@ndsu.edu.



## CRUDE OIL DEGUMMING AND REFINING

Crude vegetable oil is subjected to acid degumming for removing hydratable and non-hydratable phosphatides followed by alkali refining for removing the free fatty acids. Crude oil (1 in fig. 1) is heated to 70°C in the heater (A in fig. 1). Phosphoric acid (3) is added to the heated crude oil in the mixing tank (B) for converting the non-hydratable phosphatides to water-soluble phosphatidic acid. Soft water (4) is added to the mixing tank for the formation of gums from the hydratable phosphatides. The mixing tank outstream (5) is centrifuged (C) to separate the oil from the gums-water mix (6). Using a separator (D), the gums (7) are removed from the water (8). The degummed oil (9) is sent to a refining tank (E) maintained at 70°C for alkali refining. Sodium hydroxide solution is added for converting the free fatty acids (FFA) present in the degummed oil to oil-insoluble soaps. Proper mixing is ensured for the reaction. This is followed by the addition of wash water for dissolving the soaps, which results in the formation of soapstock that is removed from the oil using a centrifuge (F). The resulting centrifuge outstream (14) is heated to 95°C in a heater (G) and is finally sent to a vacuum oil dryer (H), with 35 mm Hg absolute pressure, to remove the remaining water in the oil. The dried degummed and refined oil is then sent to a surge tank.

The above-mentioned steps are used in conventional edible oil refining. The bleaching and deodorizing units were omitted, as these are not needed to achieve acceptable biodiesel quality.

## TRANSESTERIFICATION REACTION AND ESTER WASHING

The refined oil stream (18) enters CSTR 1 (J) maintained at 60°C, atmospheric pressure. Typically  $\geq 100\%$  excess methanol is added to the reactor along with suitable amounts of the catalyst sodium methoxide. Transesterification between triglycerides and methanol takes place in the presence of the catalyst to form methyl esters (biodiesel) and glycerol, a co-product. In addition, the remaining trace amount of free fatty acids in the refined oil reacts with sodium methoxide to form soap and methanol. The reaction products (21) are separated using decanter 1 (K) into a glycerol phase (glycerol, methanol, sodium methoxide, soaps) and ester phase (methyl esters, unreacted oil, methanol, soaps). The glycerol phase goes to a collecting tank (T). The ester phase (23) enters CSTR 2 (L), also maintained at 60°C, atmospheric pressure. A similar process occurs in CSTR 2 (L) and decanter 2 (M), using  $\geq 100\%$  excess methanol, based on the unreacted triglyceride present. The glycerol phase (27) enters the collecting tank (T), and the ester phase (28) is heated to 70°C before it goes to the ester-washing section.

Impurities in the ester phase (28), such as methanol, soaps, and free glycerol, must be separated from the methyl esters. This is done by washing the ester phase with softened water (31) at 70°C. Use of countercurrent continuous wash columns helps minimize the amount of water needed (Sheehan et al., 1998). The resultant waste stream (32) is sent to the collecting tank (T), and the washed ester stream (33) is sent to a settler tank (Q). The remaining aqueous phase is separated from the methyl esters in the settler tank (Q). The resulting ester stream (35) is then heated to 90°C before it is finally sent to the vacuum ester dryer (S), under 35 mm Hg absolute pressure, to remove the remaining traces of moisture.

## METHANOL RECOVERY AND GLYCEROL REFINING

Streams 22, 27, 32, and 34 are combined in the collecting tank (T). The resulting stream (39) is heated to the normal boiling point of methanol (64.5°C) in the heater (U). The methanol is stripped from the heated stream (40) using superheated steam (41) in the glycerol-alcohol stripper (V). The saturated methanol vapor and the steam are fed into a distillation column to recover pure methanol vapor as distillate (43). The methanol vapor (43) is condensed in a reflux condenser (X) and is recycled back (45). Bottoms (44) from the distillation column contain the condensed steam as saturated water and the remaining methanol. The hot glycerol solution (46) from the bottom of the stripper (V) is sent to a glycerol hold tank (Y). The crude glycerol (47) from this hold tank is mixed with proper amounts of HCl solution (48) in the acidulation reactor (Z). Catalyst sodium methoxide in the stream (47) reacts with HCl to form methanol and NaCl, and the soaps present in this stream react with HCl to form free fatty acids and NaCl in the acidulation reactor (Z). Using a decanter (AA), the free fatty acids and other impurities such as unreacted oil (50) are separated from the product glycerol (51).

## BIODIESEL PROCESS MODEL DEVELOPMENT

### USER-SPECIFIED INPUTS

The following parameters may be specified at the discretion of the user (fig. 2):

- Desired transesterification reaction efficiency  $\geq 85\%$ . Default value is 98% (Freedman et al., 1984; Van Gerpen et al., 2003).
- Amount of crude oil to be processed per day (kg/h). Default value is 100.
- Methanol:triglyceride mole ratio. Default value is 6 (100% excess methanol of that required by the reaction stoichiometry).
- Crude oil composition, including fatty acid profile. Soybean oil and canola oil compositions are the default values (Erickson, 1995; Hamm and Hamilton, 1999).

### PROCESS INFORMATION FOR MASS BALANCE CALCULATIONS *Crude Oil Degumming and Refining*

Mixing tank (B): Heated crude oil (2) is mixed with 0.1% of 0.85 wt% phosphoric acid solution (3) (Hernandez and Lusas, 1996) followed by the addition of soft wash water (4) equal to 75% of the phosphatide content in the crude oil (1) (Erickson, 1995).

Centrifuge (C): All phosphatide in the form of gums, all unreacted phosphoric acid, and 99.5% of stream 4 is recovered in stream 6 (Sheehan et al., 1998).

Refining tank (E): 9.5 wt% NaOH aqueous solution equal to 113% excess of that required for stoichiometric conversion of free fatty acids present in stream 9 is added. Wash water in the form of soft water equal to 15% of the mass flow rate of stream 9 is added (Sheehan et al., 1998), and 99% of the free fatty acids are converted to soaps by reaction with NaOH.

Stream 13 contains soaps formed from the reaction of free fatty acids and NaOH in the form of soapstock, triglycerides equal to 2.5 times the amount of free fatty acid loss, unsaponifiable matter associated with triglycerides in the

	A	B	C	D	E	F	G	H	I
4	Input transesterification efficiency	98							
6	Input the amount of crude oil (Kg/hr)	100							
8	Input methanol/triglyceride ratio	6							
10	Stream	1	2	3	4	5	6	7	8
11	Descr.	Crude Oil	Heater Outstream	Phosphoric acid solution	Soft water	Mixing tank Outstream	Gums-water mix	Gums	Water
14	Component								
16	Water (w)								
17	Methanol (mt)								
18	Sodium Methoxide (sm)								
19	Crude Oil								
20	TG (t)	96.00	96.00			96.00	0.48	0.48	
21	FFA (f)	0.50	0.50			0.50			
22	Phosphatides (p)	2.00	2.00			0.00			
23	Others (Unsaponifiable matter) (o)	1.50	1.50			1.50			
24	Glycerol (g)								
25	Soaps (s)								
26	Methyl Ester (me)								
27	NaOH (n)								
28	Phosphoric acid (pa)								
29	HCl (h)								
30	Gums (gu)								
31	Steam (st)								
32	NaCl (na)								
33	Total (M)	100.00	100.00	0.10	1.50	101.60	4.07	2.57	1.51
43	Fatty acid	C 16	C 16:1	C 18	C 18:1	C 18:2	C 18:3	C 20	C 22
44	Triglyceride Molecular weight (gm)	807.34	801.29	891.50	885.45	879.41	873.36	975.68	1059.83
45	Input composition of the oil(% age)	4.00	0.50	2.00	60.00	20.00	10.50	1.50	0.00
48	Molecular weight fraction	32.29	4.01	17.83	531.27	175.88	91.70	14.64	0.00
50	Molecular weight of the oil	883.43							

Figure 2. Screen view of the first eight streams in the biodiesel process model spreadsheet based on crude soybean oil.

same proportion as present in the original oil, and 99.5% of stream 11 (Sheehan et al., 1998).

Vacuum oil dryer (H): 100% moisture removal from stream 14 is achieved.

#### Transesterification and Ester Washing

CSTR 1 (J): sodium methoxide catalyst (19) equal to 1% of stream 18 (Canakci and Van Gerpen, 2001) is added in the form of a 10% solution in methanol. Methanol is added based on the user-specified methanol:triglyceride ratio and is calculated using the transesterification stoichiometry and the amount of triglyceride contained in the refined oil. Transesterification reaction efficiency of 85% is assumed regardless of the user-specified efficiency. In addition, all remaining free fatty acids in the oil react with the catalyst to form soap and methanol (Van Gerpen et al., 2003).

Stream 22 contains 60% of the methanol, all glycerol and sodium methoxide, and 10% of the soaps present in stream 21 (Van Gerpen et al., 2003).

CSTR 2 (L): sodium methoxide catalyst equal to 1% of the residual triglyceride in stream 23 is added in the form of a 10% solution in methanol (Van Gerpen et al., 2003). The amount of methanol depends on the user-specified methanol:triglyceride ratio. All the calculations are based on the transesterification stoichiometry, residual triglyceride content of stream 23, and the user-specified transesterification efficiency.

Stream 27 contains 60% of the methanol, 10% of the total amount of soaps, and all the glycerol and sodium methoxide in stream 26 (Van Gerpen et al., 2003).

Stream 30 (soft water) equals 20% of the mass flow rate of methyl esters in stream 29. As mentioned earlier, continuous countercurrent washing is assumed (Sheehan et al., 1998).

Stream 32 contains 90% of stream 30 and 100% of the methanol and soaps contained in stream 29. No ester is lost (Sheehan et al., 1998).

In stream 35, only 0.5% of stream 30 is remaining in addition to the contents of stream 29 (Sheehan et al., 1998).

Vacuum ester dryer (S): 100% moisture removal from stream 36 is achieved.

#### Methanol Recovery and Glycerol Refining

Glycerol-alcohol stripper (V): Superheated steam is at 3 mbar gauge pressure and 250°C (Hamm and Hamilton, 1999).

Stream 42 contains 100% recovery of the saturated methanol vapors and saturated steam, which flows to methanol recovery distillation column/reboiler (W).

Stream 43 contains methanol vapors with 0.05% moisture level, and stream 44 contains less than 0.5% methanol (Sheehan et al., 1998).

Stream 48 contains a 10% aqueous HCl solution equal to 50% mass flow rate of glycerine stream 47 (Sheehan et al., 1998).

Acidulation reactor (Z): Sodium methoxide reacts with HCl to form methanol and NaCl. Soaps react with HCl to form free fatty acids and NaCl.

Stream 51 contains the 80% glycerol solution obtained from the acidulation decanter (Sheehan et al., 1998).

## SAMPLE CALCULATIONS

Mass balance calculations are based on the principle of conservation of mass. Individual components are also conserved, except in process units, which involve chemical reactions. Principles of stoichiometry are used to adjust the component mass where chemical reactions occur.

Mass balance equations were applied to each individual unit for quantifying the components in each stream in the whole production process. We assumed that no net depletion or accumulation occurred and that all mass was accounted for in the flows shown in figure 1. Sample calculations follow for

three of the 27 units in figure 1: the centrifuge (C), CSTR 1 (J), and glycerol-alcohol stripper (V). These examples illustrate, respectively, a simple component balance model, a model that includes reaction stoichiometry, and a model that includes an energy balance.

### CENTRIFUGE

Stream 5 (mixing tank outstream) is separated into stream 6 (gums-water mix) and stream 9 (degummed oil). Six components were included in the centrifuge model: water (*w*), triglyceride (*t*), gums (*gu*), free fatty acids (*f*), phosphatides (*p*), and others (*o*). The mass flow rates of these components in stream 5 were calculated earlier in the model. The component mass flow rates of streams 6 and 9 are based on the process information given earlier under "Crude Oil Degumming and Refining" and are therefore calculated as follows:

Water (*w*) balances are:

$$M_{6w} = 0.995 * M_{5w} \text{ and } M_{9w} = M_{5w} - M_{6w}.$$

Triglyceride (*t*) balances are:

$$M_{6t} = 0.005 * M_{5t} \text{ and } M_{9t} = M_{5t} - M_{6t}.$$

Gums (*gu*), free fatty acids (*f*), phosphatides (*p*), and others (*o*) balances are:

$$M_{5gu} = M_{6gu} \text{ and } M_{9gu} = 0,$$

$$M_{5f} = M_{9f} \text{ and } M_{6f} = 0,$$

$$M_{5p} = M_{9p} \text{ and } M_{6p} = 0, \text{ and}$$

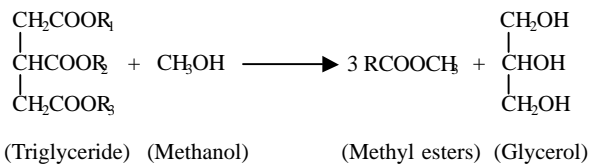
$$M_{5o} = M_{9o} \text{ and } M_{6o} = 0, \text{ respectively.}$$

where  $M_{ij}$  is the mass flow rate of any component *j* in stream *i* (kg/h). The total mass flow rate of stream 6 and 9 will each be equal to the sum of their component mass flow rates, for example:  $M_6 = M_{6w} + M_{6t} + M_{6gu}$ .

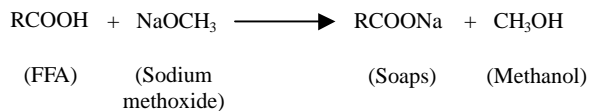
### CONTINUOUS STIRRED TANK REACTOR (CSTR) 1

Stream 18 (refined oil) is mixed with stream 19 (catalyst sodium methoxide) and stream 20 (methanol) to yield stream 21 (CSTR 1 outstream). Eight components were included in the model: triglyceride (*t*), methyl esters (*me*), glycerol (*g*), free fatty acids (*f*), methanol (*m*), sodium methoxide (*sm*), soaps (*s*), and others (*o*). The mass flow rates of these components in streams 18, 19, and 20 were calculated earlier in the model. Process information given earlier under "Transesterification and Ester Washing" was also incorporated into this model. The following two reactions take place:

1. Transesterification with 85% conversion of triglyceride:



2. Reaction between free fatty acids (FFA) in refined oil and sodium methoxide with 100% conversion of FFA to soaps:



As the above two reactions are involved in CSTR 1, the principles of stoichiometry are used to adjust the component mass flow rates in streams 19, 20, and 21. Component balances are as follows:

Triglyceride (*t*):

$$M_{21t} = M_{18t} - (0.85) * (M_{18t}) \quad (1)$$

Methyl esters (*me*):

$$M_{21me} = [(MW_{me} * 3)/(MW_t)] * (0.85 * M_{18t}) \quad (2)$$

Glycerol (*g*):

$$M_{21g} = [(MW_g)/(MW_t)] * (0.85 * M_{18t}) \quad (3)$$

Free fatty acids (*f*):

$$M_{21f} = M_{18f} - (1 * M_{18f}) = 0 \quad (4)$$

Methanol (*m*):

$$\begin{aligned}
 M_{21m} = & [M_{19m} + M_{20m}] - [(MW_m * 3)/(MW_t)] * \\
 & [0.85 * M_{18t}] + [(MW_m/MW_f)] * (1 * M_{18f}) \quad (5)
 \end{aligned}$$

Sodium methoxide (*sm*):

$$M_{21sm} = M_{19sm} - [(MW_{sm})/(MW_f)] * (1 * M_{18f}) \quad (6)$$

Soaps (*s*):

$$M_{21s} = [(MW_s)/(MW_f)] * (1 * M_{18f}) \quad (7)$$

where  $MW_j$  is the molecular weight of component *j*. Mass flow rate of the others (the unsaponifiable portion) is simply  $M_{21o} = M_{18o}$  because this component is not affected by the reactions. The total mass is conserved during the process, and the total mass flow rate of a particular stream will be equal to the sum of its component mass flow rates.

The values of  $MW_t$ ,  $MW_f$ ,  $MW_s$ , and  $MW_{me}$  were calculated as weighted averages from the vegetable oil free fatty acid composition data:

$$MW_t = \sum_k [MW_{tk} * WF_k] \quad (8)$$

$$MW_f = \sum_k [MW_k * WF_k] \quad (9)$$

where  $MW_k$  is the molecular weight of a particular free fatty acid *k*,  $MW_{tk}$  is the molecular weight of a triglyceride containing only fatty acid *k*, and  $WF_k$  is the weight fraction of fatty acid *k* in the vegetable oil.

The value of  $MW_s$  was calculated from  $MW_f$  by adding the atomic weight of sodium and subtracting the atomic weight of a hydrogen atom.

The value of  $MW_{me}$  was calculated from the transesterification stoichiometry:

$$MW_{me} = [MW_t + (MW_m * 3) + MW_g]/3 \quad (10)$$

### GLYCEROL-ALCOHOL STRIPPER

Stream 41 (superheated steam) is required to vaporize all the water and methanol from glycerol present in stream 40.

Stream 42 contains all the methanol and water from stream 40, in addition to all the steam in stream 41. All the components in stream 42 are assumed to be saturated vapor. Stream 46 is crude glycerol solution with some impurities (sodium methoxide and soaps).

Since the steam comes in direct physical contact with stream 40, an energy balance equation is needed, in addition to mass balances, to calculate the required amount of superheated steam ( $M_{41}$ ) (refer to the process information given earlier under "Methanol Recovery and Glycerol Refining"):

$$M_{40} H_{40} + M_{41} H_{41} = M_{42} H_{42} + M_{46} H_{46} \quad (11)$$

where  $H_i$  is the enthalpy of stream  $i$ . Equation 11 is rearranged to solve for  $M_{41}$ :

$$M_{41} = (M_{42} H_{42} + M_{46} H_{46} - M_{40} H_{40})/H_{41} \quad (12)$$

$M_{40}$  and all  $M_{40j}$  are known. Component balances include:

$$M_{42m} = M_{40m}, M_{42w} = M_{40w} + M_{41w}, M_{42} = M_{42m} + M_{42w}, M_{46g} = M_{40g}, M_{46sm} = M_{40sm}, M_{46r} = M_{40r}, M_{46o} = M_{40o}, \text{ and } M_{46s} = M_{40s}.$$

Other calculations necessary for solving equation 12 were performed as follows:

$$\bar{C}_{p40} [\sum_j M_{40j} C_{pj}] / [\sum_j M_{40j}] \quad (13)$$

$$H_{40} = \bar{C}_{p40} * T_{40} \quad (14)$$

$$M_{42} H_{42} = M_{42w} H_{42w} + M_{42m} H_{42m} \quad (15)$$

$$H_{42m} = C_{p,liq}(T_{40}) + L_m + C_{p,vap}(T_{42} - T_{40}) \quad (16)$$

$$M_{46} H_{46} = [\sum_j M_{46j} C_{pj}] * T_{46} \quad (17)$$

where  $M_i$  is the mass flow rate (kg/h) of any stream  $i$ ;  $\bar{C}_{p40}$  is the average heat of stream 40;  $C_{p,liq}$  and  $C_{p,vap}$  are the specific heats of liquid and vapor methanol, respectively;  $T_{40}$ ,  $T_{42}$ , and  $T_{46}$  are the temperatures of streams 40, 42, and 46, respectively;  $L_m$  is the latent heat of vaporization of methanol at its normal boiling points; and  $H_{42s}$ ,  $H_{42w}$ , and  $H_{42m}$  are the enthalpies of the components (saturated steam, water vapor, and methanol vapors, respectively) in stream 42. Enthalpies were calculated with 0°C as the reference. In the case of water and methanol vapor, saturated liquid at 0°C was the reference.  $H_{41}$  and  $H_{42w}$  were determined from superheated steam and saturated steam tables, respectively.

## DISCUSSION

Figure 2 shows the basic structure of the model inputs and outputs as seen from a portion of the spreadsheet screen. The user-specified transesterification efficiency, crude oil to be processed, and methanol: triglyceride ratio are given at the top left side of the screen. The user-specified crude oil fatty acid composition is given at the bottom of the screen. Figure 2 shows only eight of the 51 streams in the actual model, whereas all the components in the various streams are shown.

An example of the various process inputs and outputs identified by this model is listed in table 1 by choosing an

arbitrary basis of 100 kg/h crude oil entering the production plant. Outputs change according to the specified inputs. The model provides the user with a clear understanding of how the amounts of various process inputs and outputs are interrelated. Because the model is based on the compositional data of vegetable oils, it can be used for comparisons between different vegetable oils used for biodiesel production in terms of various process inputs and outputs. Table 1 shows an example of this type of comparison between soybean oil and canola oil. The process inputs and outputs are very similar for the two types of oils, despite very different fatty acid profiles. Factors that might result in large differences from the values in table 1 are a high free fatty acid content or an exceptionally low or high triglyceride molecular weight. In the case of high free fatty acid content, incorporation of acid catalysis steps is required to convert free fatty acids to methyl esters.

This article deals mainly with the mass balance calculations, and enthalpy balance calculations in the case of the glycerol-alcohol stripper (V), that were necessary for characterizing all mass flows. A separate spreadsheet has been included in the model that deals with all additional energy balance calculations for calculating the process steam requirements. A detailed description of the energy balance spreadsheet is outside the scope of the present article.

## CONCLUSION

The model described here was developed for analyzing biodiesel production. The model was based on a two-stage continuous base-catalyzed biodiesel production process. Detailed mass balance calculations have been included to quantify the flow rates of various process streams involved in the production process, according to the user-specified production level.

**Table 1. Biodiesel process inputs and outputs provided by the model with a basis of 100 kg/h crude oil processed, 98% transesterification efficiency, and 6:1 methanol:triglyceride ratio.**

	Soybean Oil (kg/h) <sup>[a]</sup>	Canola Oil (kg/h) <sup>[b]</sup>
<b>Process Inputs</b>		
Crude oil (1)	100.00	100.00
Methanol (20 + 25)	14.01	14.02
Sodium methoxide (10% solution in methanol) (19 + 24)	10.76	10.99
NaOH (9.5 wt% aqueous solution) (10)	2.58	1.59
HCl (10% aqueous solution) (48)	5.38	5.43
Process water (4 + 11 + 30)	34.91	34.74
<b>Process Outputs</b>		
Biodiesel (38)	93.52	95.30
Methanol recycled (45)	13.34	13.53
Glycerol (51)	10.31	10.41
Waste (8 + 13 + 44 + 50)	40.16	37.97
Gums (7)	2.57	1.82

<sup>[a]</sup> Composition: 96.0% triglycerides, 0.5% free fatty acids, 2.0% phosphatides, and 1.5% others (unsaponifiable matter) (Erickson, 1995). Fatty acid profile: 11% C16, 4% C18, 23% C18:1, 54% C18:2, 8% C18:3 (Erickson, 1995).

<sup>[b]</sup> Composition: 97.25% triglycerides, 0.5% free fatty acids, 1.25% phosphatides, and 1.0% others (unsaponifiable matter) (Gunstone, 2002). Fatty acid profile: 4% C16, 0.5% C16:1, 2% C18, 60% C18:1, 20% C18:2, 10.5% C18:3, 1.5% C20, 1.5% C22:1 (Hamm and Hamilton, 1999).

The model can be utilized in performing economic feasibility studies of biodiesel production in different regions simply by linking it to the economic cost data. Once a particular capacity for the biodiesel production unit has been selected (say 0.015 million m<sup>3</sup> per year), the design of the various equipment involved in the production process can be specified, based on the various stream flow rates and the desired process conditions.

#### ACKNOWLEDGEMENTS

This project is supported by the North Dakota Agricultural Experiment Station and USDA-CSREES (under Agreement No. 2003-34471-13523). The authors thank Mr. Myron Danzer, West Central Soy, Ralston, Iowa, for his biodiesel industry perspective, and Kristi Tostenson for suggestions.

#### REFERENCES

- Brown, R. C. 2003. *Biorenewable Resources*. Ames, Iowa: Iowa State University Press.
- Canakci, M., and J. Van Gerpen. 2001. A pilot plant to produce biodiesel from high free fatty acid feedstocks. ASAE Paper No. 016049. St Joseph, Mich.: ASAE.
- Erickson, D. R. 1995. *Practical Handbook of Soybean Processing and Utilization*. Champaign, Ill.: AOCS Press.
- Frazier Barnes & Associates. 2003. Mississippi statewide biodiesel feasibility study report. Memphis, Tenn.: Prepared for The Mississippi Biomass Council.
- Freedman, B., E. H. Pryde, and T. L. Mounts. 1984. Variables affecting the yields of fatty esters from transesterified vegetable oils. *J. American Oil Chemists Society* 61(10): 1638-1643.
- Freedman, B., R. O. Butterfield, and E. H. Pryde. 1996. Transesterification kinetics of soybean oil. *J. American Oil Chemists Society* 63(10): 1375-1380.
- Fryer, P. J., D. L. Pyle, and C. D. Rielly. 1997. *Chemical Engineering for the Food Industry*. New York, N.Y.: Blackie Academic and Professional.
- Gunstone, F. D. 2002. *Vegetable Oils in Food Technology: Composition, Properties and Uses*. Oxford, U.K.: Blackwell Publishing.
- Hamm, W., and R. J. Hamilton. 1999. *Edible Oil Processing*. Sheffield, U.K.: CRC Press.
- Hernandez, E., and E. W. Lusas. 1996. Practical short course in processing of vegetable oils. College Station, Texas: Texas A&M University System.
- Knothe, G., J. Krahl, and J. V. Gerpen. 2005. *The Biodiesel Handbook*. Champaign, Ill.: AOCS Press.
- Noureddini, H., and D. Zhu. 1997. Kinetics of transesterification of soybean oil. *J. American Oil Chemists Society* 74(11): 1457-1463.
- Peterson, C. L., D. L. Reece, B. L. Hammond, J. Thompson, and S. M. Beck. 1997. Processing, characterization, and performance of eight fuels from lipids. *Applied Eng. in Agric.* 13(1): 71-79.
- Sheehan, J., V. Camobreco, J. Duffield, M. Graboski, and H. Shapouri. 1998. Life cycle inventory of biodiesel and petroleum diesel for use in an urban bus. NREL/SR-580-24089. Golden, Colo.: National Renewable Energy Laboratory.
- Singh, A. P., J. C. Thompson, and B. He. 2004. A continuous-flow reactive distillation reactor for biodiesel preparation from seed oils. ASAE Paper No. 046071. St Joseph, Mich.: ASAE.
- Van Gerpen, J., D. Clemens, G. Knothe, B. Shanks, and R. Pruszko. 2003. Biodiesel production technology. Ames, Iowa: Iowa State University, Biodiesel Workshop.
- Van Wechel, T., C. R. Gustafson, and F. L. Leistritz. 2002. Economic feasibility of biodiesel production in North Dakota. Agribusiness and Applied Economics Report No. 505. Fargo, N.D.: North Dakota State University, Agricultural Experiment Station.