

## Biodiesel production from waste cooking oil: 2. Economic assessment and sensitivity analysis

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Received 18 March 2002; received in revised form 12 May 2003; accepted 13 May 2003

### Abstract

The economic feasibilities of four continuous processes to produce biodiesel, including both alkali- and acid-catalyzed processes, using waste cooking oil and the 'standard' process using virgin vegetable oil as the raw material, were assessed. Although the alkali-catalyzed process using virgin vegetable oil had the lowest fixed capital cost, the acid-catalyzed process using waste cooking oil was more economically feasible overall, providing a lower total manufacturing cost, a more attractive after-tax rate of return and a lower biodiesel break-even price. On the basis of these economic calculations, sensitivity analyses for these processes were carried out. Plant capacity and prices of feedstock oils and biodiesel were found to be the most significant factors affecting the economic viability of biodiesel manufacture.

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*Keywords:* Biodiesel; Economic assessment; After-tax rate of return; Break-even price; Sensitivity analysis

### 1. Introduction

Exploring new energy resources, such as biodiesel fuel, is of growing importance in recent years. Biodiesel, derived from vegetable oil or animal fats, is recommended for use as a substitute for petroleum-based diesel mainly because biodiesel is a renewable, domestic resource with an environmentally friendly emission profile and is readily biodegradable. The use of biodiesel as a fuel has been widely investigated. Its commercial use as a diesel substitute began in Europe in the late 1980s.

At present, the most common way to produce biodiesel is to transesterify triacylglycerols in vegetable oil or animal fats with an alcohol in the presence of an alkali or acid catalyst. Methanol is the commonly used alcohol in this process, due in part to its low cost. The products, fatty acid methyl esters (FAME), are called biodiesel and include glycerine as a by-product. Alkali-catalyzed transesterification has been most frequently used industrially, mainly due to its fast reaction rate. Sodium hydroxide or potassium hydroxide is the usual

alkali catalyst. In contrast, acid-catalyzed transesterification has received less attention because it has a relatively slow reaction rate. Nevertheless, it is insensitive to free fatty acids in feedstock oil compared to the alkali-catalyzed system. The typical acid catalyst used in the reaction is sulfuric acid.

Compared to petroleum-based diesel, the high cost of biodiesel is a major barrier to its commercialization. It costs approximately one and a half times that of petroleum-based diesel depending on feedstock oils (Prokop, 2002; Lott, 2002). It is reported that approximately 70–95% of the total biodiesel production cost arises from the cost of raw material; that is, vegetable oil or animal fats (Krawczyk, 1996; Connemann and Fischer, 1998). Therefore, the use of waste cooking oil should greatly reduce the cost of biodiesel because waste oil is available at a relatively low price.

In a previous study, four different process flowsheets for producing biodiesel from virgin vegetable oil or waste cooking oil by alkali- or acid-catalyzed transesterification were developed (Zhang et al., 2003). A comparison of these processes was presented from the point of view of their process technology. The results showed that the acid-catalyzed process from waste cooking oil was potentially a competitive alternative to the commonly used alkali-catalyzed process. Besides the

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technological evaluation, economic feasibility is also of great importance in assessing process viability. Thus, the main objective of the present article is to assess these processes on an economic basis. In this way, a better evaluation of the biodiesel production process will be achieved from both the technological and economic points of view. In addition, a sensitivity analysis of each process is presented to identify the major factors affecting the economic viability of biodiesel production.

Throughout this article, monetary values are expressed in US dollars.

## 2. Background

### 2.1. Economic studies

In previous economic studies of biodiesel production, the main economic criteria were capital cost, manufacturing cost and biodiesel break-even price. Results of three recent studies are given in Table 1. Nelson et al. (1994) evaluated the economic feasibility of a plant producing approximately 100,000 tonne/year of biodiesel. Beef tallow was transesterified with methanol in the presence of an alkali catalyst. Noordam and Withers (1996) carried out an economic study on a biodiesel plant with a capacity of approximately 7800 tonne/year biodiesel. Canola seed was used as the raw material. Canola meal produced from seed crushing was a valuable by-product, providing a credit of \$2.3 million. The estimated equipment cost for the transesterification facility was \$695,656. With respect to the transesterification unit, no detailed description of the process flowsheet or equipment sizing was provided. Other expenses, such as the costs of supervision, patents and royalties, and research and development, were not taken into consideration in their study. Bender (1999) compared seven biodiesel plants using different oilseeds (i.e., soybean seed, canola seed, sunflower seed and rapeseed) or animal fats as the raw material. The capital cost and break-even price of biodiesel for each process were de-

termined. The results for one plant are provided in Table 1. A glycerine credit was estimated based on the specification of glycerine. The overall operating costs for each process were provided, but scale-dependent expenses, such as maintenance costs, utility cost and waste disposal fees, were not described.

In the economic studies mentioned in the preceding paragraph, detailed descriptions of the processes evaluated were not provided nor were details of their economic evaluations. Also, most publications have been related to the use of various oilseeds or animal fats as the raw material. To our knowledge, no economic studies on acid-catalyzed processes using waste cooking oil as the feedstock have been reported to date. In addition, different researchers used different economic criteria to assess the biodiesel plant. Total capital cost was used by Nelson et al. (1994), whereas total biodiesel cost (i.e., total manufacturing cost) was used by Noordam and Withers (1996) to represent the economic performance of the plant. Capital equipment cost was used as an economic evaluation criterion by Bender (1999). In the present article, the economic criteria were based on fixed capital cost, total manufacturing cost, after-tax rate of return and break-even price for biodiesel.

### 2.2. Sensitivity analysis

The economic performance of a biodiesel plant (e.g., fixed capital cost, total manufacturing cost, and the break-even price of biodiesel) can be determined once certain factors are identified, such as plant capacity, process technology, raw material cost and chemical costs. However, the effects of these factors on the economic viability of the plant are also of concern. A sensitivity analysis involves measuring the relative magnitudes of these effects. This will also provide further information for the optimization of biodiesel production.

Korus et al. (1993) studied an alkali-catalyzed batch transesterification of rape oil with methanol or ethanol

Table 1  
Economic evaluations for biodiesel production plants

	Nelson et al. (1994)	Noordam and Withers (1996)	Bender (1999) <sup>a</sup>
Plant capacity	100,000 tonne/year	7800 tonne/year	10,560 tonne/year
Process type	Alkali-catalyzed continuous process	Alkali-catalyzed batch process	Alkali-catalyzed continuous process
Raw material	Beef tallow	Canola oilseed	Animal fats
Total capital cost	\$12 million	Not reported	\$3.12 million
Total manufacturing cost	\$34 million	\$5.95 million	Not reported
Biodiesel break-even price	\$340/tonne	\$763/tonne	\$420/tonne
Glycerine credit	\$6 million (\$600/tonne)	\$0.9 million (\$1450/tonne glycerine)	\$1.2 million for technical grade glycerine (\$1470/tonne); \$0.72 million for crude glycerine (\$660/tonne)

<sup>a</sup> Only results for one of the biodiesel plants evaluated is reported here.

on the laboratory scale. They concluded that the cost of feedstock oil played a significant role in determining the economic viability of the biodiesel process. Similarly, Nelson et al. (1994) commented that the significant factors affecting the cost of biodiesel were feedstock cost, plant size and glycerine by-product value. A price sensitivity analysis was performed by Noordam and Withers (1996). The variables were restricted to the costs of raw material and by-products (i.e., prices of canola seed, canola meal and glycerine). By varying the value of one variable while keeping the others unchanged, its effect on the break-even price of biodiesel was studied. Noordam and Withers (1996) reported that, on average, a \$0.01/kg increase in the canola seed cost would increase biodiesel price by \$0.03/kg. A \$0.11/kg increase in glycerine price would reduce the biodiesel price by \$0.01/kg. Because other factors, such as the ones related to process operation, were not considered, the results from such an analysis cannot be generally applied to other biodiesel plants.

On the whole, sensitivity analyses of biodiesel processes have not been used widely and only a limited number of factors have been examined. In other words, there is a lack of quantitative sensitivity analyses on the impact of a wide range of factors, such as all the possible chemical prices and variables related to operating conditions. Consequently, in the present work, sensitivity analyses of the alkali- and acid-catalyzed processes using waste cooking oil as the raw material were performed. The goal was to set up an empirical model describing the relationship between the input variables (i.e., prices of raw materials and products, plant capacity, product purities, etc.) and the output variables (i.e., economic criteria). By evaluating the sizes of parameters in the empirical model, plotting residuals and testing the model adequacy, the physical significance of each effect and the sensitivity of process economics to changes in the input factors were identified.

### 3. Process descriptions

Four different continuous processes to produce biodiesel from virgin oil or waste cooking oil were designed and simulated using HYSYS<sup>TM</sup>. Flowsheets and technological assessments of these processes were provided by Zhang et al. (2003). These processes were the focus of the economic evaluation in the present investigation. A brief description of each process follows.

Process I was an alkali-catalyzed process to produce biodiesel from virgin vegetable oil. Virgin oil and a mixture of methanol and sodium hydroxide were fed into a transesterification reactor. After the reaction (at 60 °C and 400 kPa), an effluent stream containing FAME, glycerol, methanol, unconverted oil and sodium hydroxide entered a methanol distillation column where

most of the methanol was recovered in a distillate stream. The distilled methanol, mixed with a fresh methanol stream, was recycled to the reactor. The methanol column bottom stream passed to a column for water washing to separate the FAME from glycerol, sodium hydroxide and methanol. FAME, along with unconverted oil, some water and methanol, was then forwarded to a distillation column to further remove methanol and water. From the top of that column, >99.6 wt.% (purity) FAME was obtained as a distillate. The bottom stream from the water-washing column, containing sodium hydroxide, glycerol, methanol and water, entered a neutralization reactor to remove sodium hydroxide by adding phosphoric acid. After the sodium hydroxide was removed, the stream went into a glycerine purification column where the bottom stream yielded approximately 85 or 92 wt.% glycerine as a high quality by-product depending on the extent of the distillation of methanol and water.

In process II, waste cooking oil was used as the feedstock oil to produce biodiesel in the presence of an alkali catalyst. Due to the sensitivity of the alkali-catalyzed reaction to free fatty acids found in waste oil, an esterification (i.e., pretreatment) unit was required prior to the transesterification unit to reduce the content of free fatty acid. Unrefined waste cooking oil, methanol and sulfuric acid were fed into the esterification reactor to reduce the free fatty acid content to the required level (<0.5 wt.%). The effluent stream from this reactor was sent to a liquid extraction column. Using pure glycerine as a solvent, the refined cooking oil was separated from the resulting glycerol solution of methanol, water and sulfuric acid. Then, the refined cooking oil, together with the fatty acid esters produced, entered the transesterification reactor. The bottom stream from the liquid extraction column contained methanol, glycerol, water and sulfuric acid, and proceeded to a distillation column for methanol recovery. The other parts of this process were the same as those in process I.

Process III was an acid-catalyzed process using waste cooking oil. Two transesterification reactors operating in series were required to deal with the high molar ratio (100.1) of methanol to oil. Neutralization of the acid catalyst in process III followed methanol distillation to reduce the costs of material of construction in the downstream process units. The material of construction for the transesterification and neutralization reactors and the methanol distillation column was stainless steel because of the presence of sulfuric acid. As for the other units, there was little difference between processes I and III.

Process IV was similar to process III, except that hexane was used as an extraction solvent rather than water washing to avoid the formation of emulsions. After methanol distillation, an amount of hexane, equal in volume to the amount of methanol in the feed stream, and 10 vol.% of water (based on the methanol amount)

were added to an extraction unit, where all the FAME and unconverted oil were extracted by hexane and separated from most of the glycerol, sulfuric acid and methanol. After further methanol/water washing of the hexane phase, no glycerol or sulfuric acid remained in the FAME and hexane stream. Hexane was recovered by distillation, and FAME was distilled as the biodiesel product. Other parts of the process were the same as in process III.

Zhang et al. (2003) concluded that the acid-catalyzed continuous process to produce biodiesel from waste cooking oil was a competitive alternative to the alkali-catalyzed process from a technological viewpoint. The logical next step involves economic studies and sensitivity analyses of these processes.

#### 4. Economic assessment

##### 4.1. Basis and scope of calculations

Economic evaluations were based on the following assumptions: (1) Each process was based on a plant capacity of 8000 tonne/year biodiesel. This was the same size as an existing plant in Europe (Connemann and

Fischer, 1998) and was consistent with plant sizes discussed previously (Zhang et al., 2003). In the following sensitivity analyses, two other levels of plant capacity, one at 4000 and the other at 12,000 tonne/year were considered. (2) Operating hours for the biodiesel plant were assumed to be 8000 h/year. (3) Both waste cooking oil and virgin oil, used as the feedstock for biodiesel production, are free of water and any solid impurities. The prices of the raw material oil include costs associated with impurity removal and transportation. (4) In the simulation, pump efficiency was assumed to be 70%. This was used to determine the pump shaft power. A 90% motor efficiency was used to calculate the electricity usage. Although a spare pump is usually required in a chemical plant design, no spare pumps were taken into account in this work. (5) Superheated, low and high-pressure steams were used as the heating media. Water was the cooling medium. Their specifications and prices are listed in Table 2. (6) All costs shown are in US\$. Equipment prices were updated from available mid-1996 to 2000 values using the Chemical Engineering Plant Index, where  $I_{2000} = 394$  and  $I_{1996, \text{MID}} = 382$  (Turton et al., 1998; Chemical Engineering, 2001). (7) All chemical costs including raw materials, catalysts, solvent and products are given in Table 2.

Table 2  
Basic conditions for the economic assessment of each process

Item	Specification	Price (\$/tonne) <sup>a</sup>
<i>Chemicals<sup>a</sup></i>		
Biodiesel		600
Calcium oxide		40
Glycerine	92 wt. %	1200
	85 wt. %	750
Hexane	USD	410
Methanol	99.85%	180
Phosphoric acid	Tech.	340
Sodium hydroxide	NF/FC	4000
Sulfuric acid	98%	60
Virgin canola oil		500
Waste cooking oil		200
<i>Utilities<sup>b</sup></i>		
Cooling water	400 kPa, 6 °C	\$0.007/m <sup>3</sup>
Electricity		\$0.062/kW h
Low pressure steam (superheated)	450 kPa, 210 °C	6.8
High pressure steam (superheated)	2700 kPa, 500 °C	10
<i>Waste treatment</i>		
Liquid	Hazardous	150
Solid		37
Plant capacity	8000 tonne/year biodiesel	
Thermodynamic model in process simulation	NRTL or UNIQUAC	
Methanol recovery in T-201	94%	
FAME purity	99.65 wt. %	
Methanol recovery in the pretreatment unit	94%	

<sup>a</sup> Unless specified, all prices are in US\$/tonne. Chemical prices are from National Biodiesel Board, 2000, <http://www.biodiesel.org>, Chemical Market Reporter (2000–2001), vol. 258 (22), vol. 259 (9) and <http://www.chemconnect.com>, November 2000.

<sup>b</sup> Based on the price in mid-1996 provided by Turton et al. (1998) and updated to the year 2000.

Due to the polar nature of many streams, both the non-random two liquid (NRTL) and universal quasi-chemical theory (UNIQUAC) thermodynamic/activity models were used in the process simulations. This allowed the impact of the choice of thermodynamic/activity model to be evaluated. HYSYS<sup>TM</sup> simulations of each process design provided mass and energy balances and operating conditions for each piece of equipment (Zhang et al., 2003). This information was the basis for determining the size of the process equipment in the flowsheets. Straightforward non-iterative calculations (Lin et al., 1973; Sinnott, 1999) were used for determining the equipment sizing, as discussed in detail by

Zhang (2002). Results for sizing the main equipment of each process in case 1 are shown in Table 3.

According to the definition of capital cost estimation provided by Turton et al. (1998), the economic estimation in this article is classified as a “study estimate”. It is based on the development of a process flow diagram and rough sizing of major process equipment. No further information, such as a layout plot, process instrumentation diagram or piping and instrumentation requirements, were considered. Turton et al. (1998) stated that this study estimate had a range of expected accuracy from +30% to –20%. Thus, results from such a preliminary evaluation may not accurately reflect the final

Table 3  
Equipment sizes, equipment costs and fixed capital costs for case 1 (85% glycerine byproduct using NRTL)

Type	Description	Process I	Process II	Process III	Process IV
Reactors	Esterification				
	Size, ( $D \times L$ , m)	0	0.8×2.4	0	0
	Cost, ( $\$ \times 10^{-3}$ )		80		
	Transesterification				
	Size, ( $D \times L$ , m)	1.8×5.4	1.8×5.4	2.1×6.3 <sup>a</sup>	2.1×6.3 <sup>a</sup>
	Cost, ( $\$ \times 10^{-3}$ )	290	290	673 <sup>a</sup>	673 <sup>a</sup>
Neutralization	Size, ( $D \times L$ , m)	0.3×1	0.3×1	0.5×1.5	0.4×1.2
	Cost, ( $\$ \times 10^{-3}$ )	21	21	38	29
Columns	Methanol distillation				
	Size, ( $D \times H$ , m)	0.6×10	1×12	1×10	1×10
	Cost, <sup>b</sup> ( $\$ \times 10^{-3}$ )	140	0.6×10 357 140	324	324
	Washing column				
	Size, ( $D \times H$ , m)	0.8×10	1×10	1×10	1×10
	Cost, ( $\$ \times 10^{-3}$ )	100	0.8×10 240 100	109	0.8×10 240 220
	FAME distillation				
	Size, ( $D \times H$ , m)	1.2×12	1.2×12	1×12	1.5×16 <sup>c</sup>
	Cost, <sup>b</sup> ( $\$ \times 10^{-3}$ )	157	157	168	256 <sup>c</sup>
	Glycerine purification				
Size, ( $D \times H$ , m)	N/A	0.5×10	0.6×10	0.8×10	
Cost, <sup>b</sup> ( $\$ \times 10^{-3}$ )		106	135	145	
Heat exchangers, cost ( $\$ \times 10^{-3}$ )	4	16	12	12	
Pumps, cost ( $\$ \times 10^{-3}$ )	45	72	40	49	
Others (separator, vacuum system), cost ( $\$ \times 10^{-3}$ )	46	57	73	42	
Total basic module cost ( $\$ \times 10^{-6}$ ), $C_{BM0}$	0.61	1.10	1.01	1.19	
Total bare module cost ( $\$ \times 10^{-6}$ ), $C_{BM}$	0.81	1.64	1.57	1.99	
Contingency fee ( $\$ \times 10^{-6}$ ), $C_{CF} = 0.18C_{BM}$	0.14	0.29	0.28	0.36	
Total module cost ( $\$ \times 10^{-6}$ ), $C_{TM} = C_{BM} + C_{CF}$	0.95	1.94	1.85	2.35	
Auxiliary facility cost ( $\$ \times 10^{-6}$ ), $C_{AC} = 0.3C_{BM0}$	0.22	0.39	0.36	0.42	
Fixed capital cost, $C_{FC}$ ( $\$ \times 10^{-6}$ ), $C_{FC} = C_{TM} + C_{AC}$	1.17	2.33	2.21	2.77	
Working capital ( $\$ \times 10^{-6}$ ), $C_{WC} = 0.15C_{FC}$	0.17	0.35	0.33	0.42	
Total capital investment ( $\$ \times 10^{-6}$ ), $C_{TC} = C_{FC} + C_{WC}$	1.34	2.68	2.55	3.19	

<sup>a</sup> Includes two reactors operated in series.

<sup>b</sup> Includes the overhead condenser, recycle pump and bottom reboiler.

<sup>c</sup> Includes hexane distillation.

profitability of a chemical plant but can be used as a tool for comparison of several process alternatives (Turton et al., 1998).

In the present study, “economic assessment” refers to the evaluation of fixed capital cost, total manufacturing cost, after-tax rate of return, and break-even price of biodiesel. Fixed capital cost represents the cost of constructing a new plant (also called grass-roots capital cost). Generally, fixed capital cost consists of three parts: total bare module capital cost, contingencies and fees, and costs associated with auxiliary facilities. Total bare module capital cost is the sum of the cost of each piece of equipment in the process. Contingencies and fees are defined as a fraction of the total bare module capital cost (e.g., 18% was used in the present study) to cover unforeseen circumstances and contractor fees (Ulrich, 1984; Turton et al., 1998). Expenses of auxiliary facilities include items such as the purchase of land, installation of electrical and water systems and construction of all internal roads. They are usually represented by 30% of the total basic module cost (i.e., the sum of total bare module capital cost and contingency and fees) (Ulrich, 1984; Turton et al., 1998). Total capital investment is calculated by adding the fixed capital cost to the working capital cost. The latter is usually a fraction of the fixed capital cost (e.g., 15% was used in this work) (Ulrich, 1984). Fixed capital costs and total capital investments for each process are presented in Table 3.

Total manufacturing cost refers to the cost of the day-to-day operation of a chemical plant and is usually divided into three categories: direct manufacturing costs, indirect manufacturing costs (or fixed manufacturing cost) and general expenses (Ulrich, 1984; Turton et al., 1998). Direct manufacturing costs consist of raw material costs, catalyst and solvent costs, operating labor fees, supervisory and clerical labor fees, utilities (including waste disposal), maintenance and repairs, operating supplies, laboratory charges, and expenses for patents and royalties. In brief, all charges related to materials and labor belong to this category. In the present study, the costs of raw material, catalyst and solvent, and utilities were calculated based on the price of each chemical (Table 2), and their flowrate and usage from the HYSYS<sup>TM</sup> simulation results (Zhang et al., 2003). The operating labor fee was obtained from operator requirements for various pieces of process equipment (Turton et al., 1998). It was assumed that an operator worked on average 49 weeks/year and there were three shifts a day for a continuously running plant. Operator salary was estimated at \$47,850/year (\$24/h). Other expenses, such as supervisory and clerical labor fees, maintenance and repair expenses, and operating supplies charges were calculated individually and multiplied by related factors, as shown in Table 4. Indirect manufacturing costs include overhead, packaging, storage, local taxes, insurance and depreciation. All of the

items in this category are independent of the production rate in a plant. The last category, general expenses, includes administrative costs, distribution and selling costs, and research and development charges. Similarly, items pertaining to indirect manufacturing costs and general expenses were also computed and multiplied by various constant factors, which are commonly applied to economic assessments and are shown in Table 4 (Ulrich, 1984; Turton et al., 1998). Glycerine cost was regarded as a by-product credit.

After-tax rate of return is a general economic performance criterion for the preliminary evaluation of a chemical plant and is defined as the percentage of net annual profit after taxes relative to the total capital investment. Net annual profit after taxes ( $A_{NNP}$ ) is equal to income after taxes ( $A_{IT}$ ) and is half of the net annual profit ( $A_{NP}$ ) when a 50% corporate tax rate is used (Ulrich, 1984). After-tax rate of return was also chosen as the response variable and objective function in the economic assessment and sensitivity analyses of this work.

Break-even price is defined as the price for which revenue from biodiesel product is the same as total manufacturing cost of a plant. Although the break-even price for biodiesel has been quoted as an economic criterion in previous publications (Nelson et al., 1994; Noordam and Withers, 1996; Bender, 1999), it underestimates the actual commercial cost of biodiesel by excluding plant profit and taxes.

## 4.2. Results and discussion

Four cases were considered: case 1 was for 85 wt.% glycerine using NRTL; case 2 was 85 wt.% glycerine using UNIQUAC; case 3 was 92 wt.% glycerine using NRTL; case 4 was 92 wt.% glycerine using UNIQUAC. Following the procedures given in the previous section, the values of fixed capital cost, total manufacturing cost and after-tax rate of return for cases 1 through 4 in each process were determined. Based on the results it appears that differences in fixed capital cost, total manufacturing cost or after-tax rate of return between each case were insignificant. Thus, only detailed results for case 1 (85 wt.% glycerol and NRTL) in each process are reported. A summary of the economic performance of each process for case 1 is shown in Fig. 1 and Tables 3–5.

### 4.2.1. Fixed capital cost

As shown in Table 3, the cost of the transesterification reactor(s) forms a significant part of the capital cost, especially for acid-catalyzed processes III and IV. The costs associated with the transesterification unit amounted to 35–46% of the total bare module cost in the acid-catalyzed processes (Table 5). Large amounts of methanol required in the acid-catalyzed process led to an increased reaction volume. In addition, the presence

Table 4  
Total manufacturing cost and after-tax rate of return for case 1 ( $\$ \times 10^{-6}$ )

	Process I	Process II	Process III	Process IV
<i>Direct manufacturing cost</i>				
Oil feedstock	4.20	1.68	1.65	1.65
Methanol	0.17	0.18	0.31	0.31
Catalyst and solvent	0.32	1.41	0.07	0.08
Operating labor	0.58	1.03	0.85	0.92
Supervisory and clerical labor, 15% of operating labor	0.09	0.15	0.13	0.14
<i>Utilities</i>				
LP steam	0.03	0.06	0.17	0.17
HP steam	0.06	0.05	0.06	0.07
Electricity	0.02	0.02	0.03	0.03
Cooling water	0.005	0.008	0.02	0.04
<i>Waste disposal</i>				
Liquid	0.009	0.28	0.28	0.27
Solid	0.004	0.004	0.06	0.06
Maintenance and repairs, 6% of $C_{FC}$	0.07	0.14	0.13	0.17
Operating supplies, 15% of maintenance and repairs	0.01	0.02	0.02	0.02
Laboratory charges, 15% of operating labors	0.09	0.15	0.13	0.14
Patents and royalties, 3% of total manufacturing cost	0.20	0.21	0.15	0.17
Subtotal, $A_{DME}$	5.14	4.75	3.28	3.52
<i>Indirect manufacturing cost</i>				
Overhead, packaging and storage, 60% of the sum of operating labor, supervision and maintenance	0.44	0.79	0.66	0.74
Local taxes, 1.5% of $C_{FC}$	0.02	0.03	0.03	0.04
Insurance, 0.5% of $C_{FC}$	0.006	0.01	0.01	0.01
Subtotal, $A_{IME}$	0.46	0.83	0.71	0.79
Depreciation, $A_{BD}$ , 10% of $C_{FC}$	0.12	0.23	0.22	0.28
<i>General expenses</i>				
Administrative costs, 25% of overhead	0.11	0.20	0.17	0.18
Distribution and selling cost, 10% of total manufacturing cost	0.68	0.71	0.52	0.56
Research and development, 5% of total manufacturing cost	0.34	0.35	0.26	0.28
Subtotal	1.14	1.26	0.94	1.03
Total production cost	7.59	7.76	5.92	6.35
Glycerine credit	0.73	0.68	0.77	0.73
Total manufacturing cost, $A_{TE}$	6.86	7.08	5.15	5.62
Revenue from biodiesel	4.80	4.80	4.80	4.80
Net annual profit, $A_{NP}$	-2.06	-2.28	-0.35	-0.82
Income taxes, $A_{IT}$ , 50% of $A_{NP}$	-1.03	-1.14	-0.18	-0.41
Net annual profit after taxes, $A_{NNP}$	-1.03	-1.14	-0.18	-0.41
After-tax rate of return, $I = [A_{NNP} - A_{BD}]/C_{TC}$ , (%)	-85.27	-51.18	-15.63	-21.48

of sulfuric acid required the construction material of the reactor to be stainless steel. Consequently, the reactor sizes and costs in processes III and IV were much higher than those in processes I and II.

The costs related to the separation units in process IV were much higher than in the other processes (Table 5). This is because the addition of hexane and methanol/water increases the loads in the washing columns, FAME distillation and glycerine purification columns in process IV. This in turn, requires larger sizes and higher costs for these units. The cost of the separation units in process III is also high because the presence of sulfuric acid in some units requires stainless steel construction

(e.g., the methanol distillation column). The pretreatment unit contributed significant unique costs, resulting in a large increase in the fixed capital cost in process II.

Overall, process I, the alkali-catalyzed process using virgin oil as the raw material, had the lowest fixed capital cost. The fast transesterification reaction reduced the size and cost of the reactor. The use of virgin oil with low-levels of free fatty acids removed the need for a pretreatment unit. The use of less corrosive sodium hydroxide reduced the requirement for stainless steel as the material of construction in some pieces of process equipment. Consequently, the fixed capital cost and the total capital investment in process I were much lower

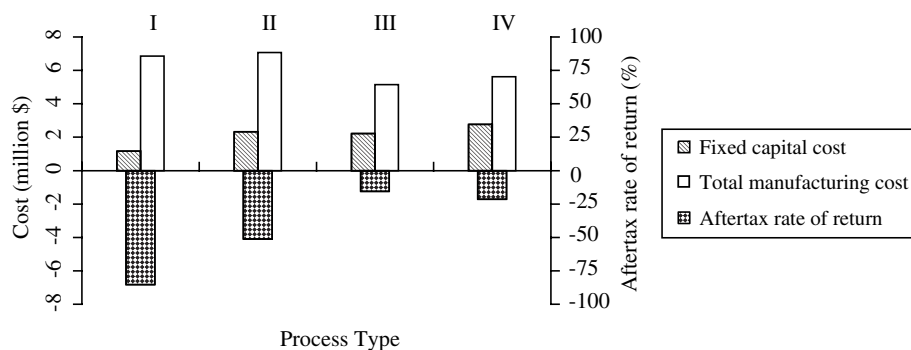


Fig. 1. Economic performance of processes I, II, III, and IV for case 1.

Table 5  
Summary of economic criteria for each process for case 1

	Process I	Process II	Process III	Process IV
<i>Fixed capital cost</i>				
Pretreatment unit				
Cost (\$ × 10 <sup>-6</sup> )	0	0.76	0	0
Percentage (%)	0	46	0	0
Transesterification unit				
Cost (\$ × 10 <sup>-6</sup> )	0.33	0.31	0.72	0.72
Percentage (%)	41	19	46	36
Separation unit				
Cost (\$ × 10 <sup>-6</sup> )	0.48	0.57	0.85	1.27
Percentage (%)	59	35	54	64
Total bare module cost				
Cost (\$ × 10 <sup>-6</sup> )	0.81	1.64	1.57	1.99
Percentage (%)	100	100	100	100
<i>Total manufacturing cost</i>				
Direct manufacturing				
Cost (\$ × 10 <sup>-6</sup> )	5.14	4.75	3.28	3.52
Percentage (%)	75	67	64	63
Indirect manufacturing				
Cost (\$ × 10 <sup>-6</sup> )	0.46	0.83	0.71	0.79
Percentage (%)	7	12	14	14
General expenses				
Cost (\$ × 10 <sup>-6</sup> )	1.14	1.26	0.94	1.03
Percentage (%)	17	18	18	18
Depreciation				
Cost (\$ × 10 <sup>-6</sup> )	0.12	0.23	0.22	0.28
Percentage	2	3	4	5
Subtotal				
Cost (\$ × 10 <sup>-6</sup> )	6.86	7.08	5.15	5.62
Percentage	100	100	100	100
<i>Profitability</i>				
Net annual profit after taxes (\$ × 10 <sup>-6</sup> )	-2.06	-2.28	-0.35	-0.82
After-tax rate of return (%)	-85.27	-51.18	-15.63	-21.48
Break-even price of biodiesel (\$/tonne)	857	884	644	702

than those in the other processes, as shown in Fig. 1 and Table 3.

As discussed in Section 2.1, previous economic studies of biodiesel plants lacked descriptions of process

equipment sizing. Thus, a detailed comparison of our results with those of others can not be carried out. According to the results provided by Noordam and Withers (1996), their equipment cost for the transesterification unit with a biodiesel production capacity of 7800 tonne/year was estimated to be \$0.70 million. If this value is updated to the cost in the year 2000, it is approximately \$0.72 million. In this work, the bare module cost in process I with a size of 8000 tonne/year biodiesel was \$0.81 million. Compared to the \$3.12 million capital cost of a 10,000 tonne/year biodiesel production process from animal fats given by Bender (1999), our alkali-catalyzed process II of 8000 tonne/year biodiesel from waste cooking oil, had a total capital investment of \$2.68 million. Regardless of the differences in the plant capacities, our results appear reasonable.

#### 4.2.2. Total manufacturing cost

Results for the total manufacturing costs of processes I through IV are shown in Tables 4 and 5. The direct manufacturing cost represents approximately 63–75% of the total manufacturing cost in each process and should be the focus of efforts to minimize the total manufacturing cost. The cost of feedstock oil comprises up to 50% of the direct manufacturing cost for the acid-catalyzed processes (III and IV) (Table 4). Using virgin oil as the raw material in process I results in approximately 80% of the direct manufacturing cost being related to the raw material cost. In process II, despite the fact that the addition of a pretreatment unit has increased direct manufacturing costs associated with solvent and operating labor, the raw material cost is approximately 35% of the total direct manufacturing cost. This is still the largest component of the total direct manufacturing cost. Consequently, raw material costs appear to play the most important role in direct manufacturing costs, which in turn substantially affect the total manufacturing cost.

Although process I has the lowest value for equipment costs (\$0.81 million) and total capital investment (\$1.34 million) (Table 3), the use of virgin vegetable oil greatly increases the raw material cost (\$4.2 million). In contrast, using waste cooking oil can reduce the raw material cost to \$1.68 million in alkali-catalyzed process II (Table 4). However, due to the sensitivity of the alkali-catalyzed process to free fatty acids, the cost associated with the addition of a pretreatment unit accounts for approximately 46% of the total bare module cost (Table 5). Moreover, the increased solvent cost in this unit negates the by-product glycerine credit (Table 4).

Noordam and Withers (1996) concluded that the total manufacturing cost, updated to the year 2000, for their plant using canola seed was \$6.1 million. If this value is assumed to cover the same range as the total manufacturing cost in the present study, the total

manufacturing cost of process I (i.e., \$6.86 million) is similar to their result. This indicates that the use of oilseed as the raw material can lower the raw material cost but will not have much influence on the total manufacturing cost of the plant. The net cost associated with seed procurement and oil extraction would be \$4.61 million in their study, which completely offsets the raw material cost of virgin oil in process I (\$4.2 million).

The credit for the glycerine by-product has a significant impact on the net value of the total manufacturing cost. As shown in Table 4, the glycerine credit is up to \$0.70 million at a rate of \$0.75/kg for 85 wt.% purity in each process. Bender (1999) reported a \$0.72 million credit for crude glycerine, which is similar to our results. If the glycerine credit is not considered in process II, the total production cost and total manufacturing cost will be somewhat reduced to \$7.5 million because of the absence of the glycerol purification unit. This value is still higher than the value of \$7.08 million with the glycerine credit (Tables 4 and 5). Moreover, the possible increase in waste disposal fees has not yet been taken into account. In other words, the glycerine credit can lead to a reduction in total production costs by 6% in process II. Using a similar calculation in absence of glycerine credit, a reduction of approximately 13–14% in the total production cost will be achieved in processes III and IV.

#### 4.2.3. After-tax rate of return and break-even price

Each process appears to have a negative net annual profit, as well as a negative after-tax rate of return (i.e., after-tax rate of loss) (Tables 4 and 5). This suggests that suitable government subsidies may be required to produce positive rates of return. As discussed earlier, the absolute values of these numbers possess uncertainties arising from the HYSYS<sup>TM</sup> simulation and the preliminary nature of the economic estimation methods used in the present study. Nevertheless, the relative values or the differences among these results are of importance. Compared to the other processes, process I has the highest after-tax rate of loss. This is because the net annual loss produces a large value when divided by the relatively small fixed capital cost for process I. The after-tax rate of loss amounts to 85% of the total capital investment in process I (Table 4). Overall, the acid-catalyzed processes III and IV appear to be more economical than the alkali-catalyzed processes I and II because of the lower production costs.

The break-even prices of biodiesel from all processes are listed in Table 5 and range from \$644 to \$884/tonne. The acid-catalyzed processes III and IV have lower break-even prices than processes I and II. For a similar biodiesel plant, Noordam and Withers (1996) estimated the break-even price for biodiesel from canola oilseed by alkali catalysis at \$763/tonne, in close agreement with the range in our results.

An alternative acid-catalyzed process to achieve 99% oil conversion to FAME using a 245:1 molar ratio of methanol to oil, 80 °C and 3.5 mol% (based on the whole mixture) sulfuric acid was discussed by Zhang et al. (2003). Although the amount of unconverted oil waste was reduced as a result of increasing the oil conversion, the large amount of excess methanol led to a large increase in the sizes of the methanol distillation and subsequent separation units, as well as an increase in utility costs. Economic evaluation results for this process showed that a  $\$3.24 \times 10^6$  fixed capital cost, a  $\$8.04 \times 10^6$  total manufacturing cost, a  $-52\%$  after-tax rate of return and a  $\$1005/\text{tonne}$  break-even price for biodiesel were required. Therefore, this process is not economically reasonable even though it is technologically feasible.

## 5. Sensitivity analysis

After completion of the economic assessment of the biodiesel production processes, sensitivity analyses for these processes were conducted to determine the sensitivity of the after-tax rate of return to changes in a variety of factors thought to plausibly significantly impact the process economics. As an example for discussion, process III, the acid-catalyzed process to produce biodiesel from waste cooking oil, was used. Detailed procedures have been presented elsewhere (Zhang, 2002).

Nine factors, identified in Table 6, were investigated. The purity of glycerine was treated as a block factor.

One block (block I) used a 85 wt.% glycerine purity, while the other (block II) used a 92 wt.% glycerine purity. Each block had different price levels corresponding to the glycerine purity. A  $2_{IV}^{9-4}$  fractional factorial design, involving 32 simulation runs, was used in the present study to determine the physical significance of each factor on the after-tax rate of return (Box et al., 1978). A polynomial model was found to describe the relationship of each factor with the response variable. The physical significance of each effect was determined based on the relative magnitudes of the parameter estimates in the model and plots of residuals (i.e., discrepancies between model predictions and simulation results) and of the sum of squares of residuals with increasing numbers of factors added to the model. Results showed that plant capacity  $X_1$ , oil price  $X_2$  and biodiesel price  $X_6$  had the greatest effect on the after-tax rate of return. Similar analyses were applied to processes II and IV with similar results (Table 7).

The sensitivities of the break-even price of biodiesel to the prices of waste oil and glycerine for a plant producing 8000 tonne/year biodiesel were also calculated. The results showed that an increase of  $\$0.01/\text{kg}$  in waste oil price would cause an increase of  $\$0.02/\text{kg}$  in the biodiesel break-even price. However, a reduction of  $\$0.01/\text{kg}$  in the biodiesel break-even price can be obtained by an increase of  $\$0.06/\text{kg}$  in the price of crude glycerine (85 wt.%). These results suggested that the influence of the oil price on the biodiesel price is larger than the influence of the glycerine price. Comparing

Table 6  
Descriptions of factors in sensitivity analysis for process III

Factor	Description	Level	
		-1	1
$X_1$	Plant capacity	4000 tonne/year	12,000 tonne/year
$X_2$	Price of waste cooking oil	\\$100/tonne	\\$300/tonne
$X_3$	Methanol price	\\$80/tonne	\\$280/tonne
$X_4$	Glycerine price (85 wt.%, block I)	\\$650/tonne	\\$850/tonne
	Glycerine price (92 wt.%, block II)	\\$1100/tonne	\\$1300/tonne
$X_5$	Thermodynamic model	NRTL	UNIQUAC
$X_6$	Biodiesel price	\\$400/tonne	\\$600/tonne
$X_7$	Price of sulfuric acid	\\$50/tonne	\\$250/tonne
$X_8$	Methanol recovery in T-201	90%	98%
$X_9$	Final FAME (biodiesel) purity	99.6 wt.%	99.7 wt.%

Table 7  
Results of sensitivity analyses<sup>a</sup>

Process	Process description	Major effects	Minor effects
Process II	Alkali-catalyzed process using sodium hydroxide catalyst	Plant capacity, $X_1$ ; price of waste cooking oil, $X_2$ , and biodiesel price, $X_6$	Interaction of $X_1X_6$ , interaction of $X_1X_2$
Process III	Acid-catalyzed process using sulfuric acid catalyst	Plant capacity, $X_1$ , price of waste cooking oil, $X_2$ , and biodiesel price, $X_6$	Interaction of $X_1X_6$ , interaction of $X_1X_2$ and thermodynamic model $X_5$
Process IV	Acid-catalyzed process using hexane extraction	Plant capacity, $X_1$ ; price of waste cooking oil, $X_2$ and biodiesel price, $X_6$	Interaction of $X_1X_2$ , methanol recovery, $X_8$ , and interaction of $X_1X_6$

<sup>a</sup> The sensitivity analysis of process I was not carried out.

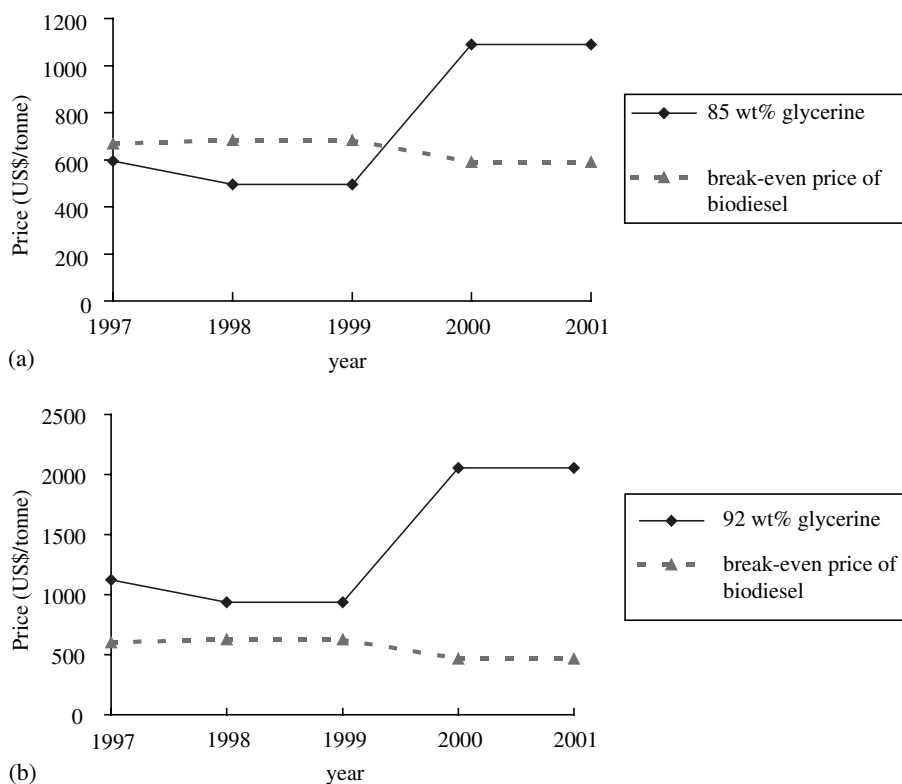


Fig. 2. Yearly change in price for glycerine and its impact on the break-even price of biodiesel.

these values to the results of Noordam and Withers (1996) indicates no practical difference. However, the fluctuation of the glycerine price may require additional consideration. Based on the prices of high-purity glycerine (>99 wt.%) in the Chemical Market Reporter (1997–2001), the price for 92 wt.% glycerine was assumed to be 85% of the high-purity price. The price of 85 wt.% glycerine was assumed to be 45% of the price of 92 wt.% glycerine (Sonntag, 1991; Bender, 1999). With the fluctuation in the glycerine price, the expected change in the break-even price of biodiesel is shown in Fig. 2a (85 wt.% glycerine) and b (92 wt.% glycerine). These results indicate that, despite the seemingly large fluctuation in glycerine price, it has a relatively small impact on the biodiesel break-even price.

## 6. Conclusions

On the basis of the economic assessment of four continuous alkali- and acid-catalyzed processes using virgin oil or waste cooking oil as the raw material, the following conclusions were made. The alkali-catalyzed process using virgin oil (process I) had the lowest total capital investment because of the relatively small sizes and carbon steel construction of most of the process equipment. For a plant producing 8000 tonne/year biodiesel, the total capital investment in process I was

approximately half of that in the other processes. Process I thus requires the least initial investment when building a biodiesel plant.

Raw material costs account for a major portion of the total manufacturing cost. Thus, reduction of the raw material cost should be the first step in optimizing the total manufacturing cost. Virgin oil costs approximately 2–3 times more than waste cooking oil indicating that use of virgin oil leads to a substantial increase in total manufacturing cost. As a result, although process I had the lowest cost requirement for building a biodiesel plant, it had a high manufacturing cost offsetting any economic advantage in terms of return on investment or biodiesel break-even price.

When waste oil of low cost was used as the raw material, alkali-catalyzed process II required a pretreatment unit to reduce the content of free fatty acids. The cost associated with this pretreatment unit, including the cost for extra solvent, more than balanced the credit of using waste oil. This led to a reduced economic feasibility for process II. The acid-catalyzed system was insensitive to free fatty acids and no pretreatment unit was required. Accordingly, the acid-catalyzed processes to produce biodiesel from waste cooking oil (processes III and IV) had lower total manufacturing costs than the alkali-catalyzed process, so that the acid-catalyzed process would cost less to operate. Glycerine was a valuable by-product, which could add an appreciable credit to

reduce the total manufacturing cost by approximately 10% for a plant with 8000 tonne/year biodiesel capacity. Based on after-tax rate of return and break-even price of biodiesel, the acid-catalyzed processes III and IV were economically competitive alternatives to the alkali process for biodiesel production. Sensitivity analyses of different processes for biodiesel production showed that plant capacity, the price of waste cooking oil and the price of biodiesel were the major factors affecting the economic feasibility of the biodiesel production.

### Acknowledgements

The Natural Sciences and Engineering Research Council of Canada supported this work. The authors acknowledge Dr. A.Y. Tremblay for his assistance with the HYSYS™ simulation and Dr. B.C.-Y. Lu for his assistance with the thermodynamic model.

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