

**EVALUATING THE ECONOMIC FEASIBILITY OF CANOLA
BIODIESEL PRODUCTION IN NORTH DAKOTA**

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Title

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ABSTRACT

Tapasvi, Dhruv; M.S.; Departments of Agricultural and Biosystems Engineering, and Agribusiness and Applied Economics; College of Graduate and Interdisciplinary Studies; North Dakota State University; April 2006. Evaluating the Economic Feasibility of Canola Biodiesel Production in North Dakota. Major Professors: Dr. Dennis Wiesenborn and Dr. Cole Gustafson.

Numerous factors have pushed energy from biomass to the forefront of policy and industry discussions. Large harvests of traditional crops, low farm prices, dependence on foreign energy sources, and environmental problems have increased interest in renewable energy sources. Tools are needed to evaluate and compare different available feedstocks and to identify parameters and modifications for the production of renewable fuels such as biodiesel.

The first paper examines the development of a biodiesel process model using commonly available 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 is modeled as 27 units with 51 flows and 18 components. Mass flow rates and compositions of the process input and output streams are 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 are 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 inputs of 13.8, 10.8, and 34.7 (in kg/h) for methanol, 10% sodium methoxide in methanol, and process water, respectively; and outputs of 93.5, 10.3, and 55.6 for soy biodiesel, glycerol, and waste stream, respectively.

In the second paper, the mass flow rate data from the developed biodiesel process model are linked to cost data for evaluating the economic feasibility of biodiesel production in North Dakota with canola oil as the feedstock. Estimations of capital investment cost and total annual biodiesel product cost are conducted for two canola biodiesel production plants with 5 and 30 million gallons per year (MGY) capacities. These capacities were selected based on North Dakota and neighboring states' biodiesel demands, respectively. Capital investment cost analysis shows the presence of considerable economies of scale for the biodiesel production process for the two capacities. These cost calculations are based on the purchased equipment cost calculated from the equipment specifications. Total annual biodiesel product cost analysis shows that the major portion (>80%) of the total product cost is the raw material cost, similar to the analysis of previous economic feasibility studies. Cost benefits from the economies of scale are still present for the fixed charges, general expenses, and the manufacturing costs (other than the raw material costs) in the annual product cost calculations for the two production plant capacities. Finally, based on the gross profit evaluation for both plants, this study concludes that it is more worthwhile to invest in the 30 MGY production plant because of the greater cost returns from the economies of scale benefits. The results are more encouraging after the incorporation of the federal biodiesel tax incentive and favor the investment for biodiesel production in North Dakota.

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GENERAL INTRODUCTION

Biodiesel is a diesel fuel substitute produced from domestic renewable resources such as vegetable oils, animal fats, and recyclable cooking oils. Chemically, biodiesel is defined as the alkyl esters of long chain fatty acids derived from renewable lipid sources. Primarily, it has been processed commercially in the European Union and United States. In the United States, biodiesel has been registered with the United States Environmental Protection Agency as a pure fuel or as a fuel additive and is a legal fuel for commerce (Knothe et al., 2005).

Biodiesel has only become a commercial reality in the last two decades, although the idea of vegetable oil-based fuels has been around since the turn of the twentieth century. In 1912, Rudolf Diesel wrote, "The use of vegetable oils for the engine fuels may seem insignificant today but such oils may become in the course of time as important as petroleum and the coal tar products of the present time" (Page 118, Quick, 1989). Historically, agricultural producers conducted fieldwork with draft animals. They would plant a portion of their land to feed their "horsepower." The use of biodiesel as an energy for compression-ignition engines may allow producers of the modern world to return to the idea of "growing" their fuel.

Researchers have examined the effects of vegetable oils on long-term diesel engine performance. In 1985, Ziejewski of North Dakota State University, Fargo, ND, in his Ph.D. research, tested the effects of different blends of sunflower oil and safflower oil with regular diesel on long-term diesel engine performance. Ultimately, a large amount of testing confirmed that these blends produce excessive carbon deposits in engines.

Researchers determined that vegetable oils must be chemically altered before being used as alternative fuels.

Biodiesel can be used as B 100 (neat) or in a blend with petroleum diesel. A blend of 20% biodiesel with 80% petro-diesel, by volume, is termed “B 20.” A blend of 2% biodiesel with 98% petro-diesel is “B 2.”

Biodiesel demand drivers

Energy security

Biodiesel will decrease the United States’ dependence on imported petroleum. The United States uses approximately 20 million barrels of oil a day (that’s about 3 gallons a day for every person in the country), more than half of which is imported. In 2005, just over 68% of crude oil consumed in the United States was imported. In fact, United States crude oil production has dropped steadily since 1988 (U.S. Department of Energy/EIA, 11/2005). Domestic production of an alternative fuel such as biodiesel can provide the required energy security. Incrementally increasing the biofuels content of motor vehicle fuel (gasoline and diesel) from 1.2 to 4.0% between 2002 and 2016 would displace a total of 2.9 billion barrels of crude oil (Urbanchuk, 2001).

Emission benefits

Biodiesel and biodiesel blends with petro-diesel generate reductions in all regulated emissions except nitrogen oxides, which can also be effectively eliminated with the use of normal mechanical remediation techniques (e.g. catalysts or engine timing changes). Biodiesel has unique combustion qualities as an oxygenated plant/animal based fuel, as it retains oxygen present in the original oil or fat source—giving it a definite edge over regular petro-diesel with respect to emissions. Also, as it is derived from natural sources,

biodiesel does not contain any aromatic hydrocarbons, metal or crude oil residues (trace amounts of sulfur can occur if the oil or fat source contains some protein contaminants). Also, the sulfur content of biodiesel is either nil or negligible depending upon the feedstock used for its production. Studies have shown that using B20 yields percent reductions of 16-33% in total particulates, 11-25% in carbon monoxide, and 19-32% in total hydrocarbons. Moreover, emissions of polyaromatic hydrocarbons are also lower for biodiesel-blended fuels when compared with regular diesel (Koo-Oshima et al., 1998).

Biodegradability

Biodiesel also has desirable degradation attributes. Studies conducted at the University of Idaho tested the biodegradability of biodiesel in aquatic environments and showed that biodiesel degrades more rapidly than a test sugar, dextrose. Biodiesel samples were 95% degraded at the end of 28 days where as the regular petro-diesel was approximately 40% degraded after 28 days (Zhang et al., 1995).

Enhanced lubricity

Another advantage of biodiesel over petro-diesel is its superior lubricity. In fact, refined petro-diesel should contain sulfur (a major pollutant) to help with lubricity. In the near future, government regulations will require improved refinery processes that will reduce the sulfur content in diesel. These mandated reductions in allowable sulfur content are likely to go into effect on September 1, 2006 and these will help increase demand for biodiesel (Coltrain, 2002; Chevron corporation, 2006).

Market for excess feedstock

Biodiesel also provides a market for excess production of vegetable oils and animal fats. There is increasing demand around the world for soybean meal, which is a co-product

of soybean oil extraction, to provide protein for human and animal consumption. If new markets are not found for soybean oil, then prices will be low and farmers will have even more difficulty producing a profit.

Reasons for the study

The biodiesel industry in the United States is poised for rapid expansion and is now emerging as a key player in United States energy and agricultural policy. A number of current legislative initiatives exist that could significantly increase the country's biodiesel demand. For example, on August 8, 2005, President Bush signed the Energy Policy Act (EPAAct) of 2005 with biodiesel provisions. The legislation sets a federal goal for increased national use of renewable fuels such as biodiesel. Under the measure, renewable fuels would account for 5 billion gallons of national use by 2012 (US Department of Energy, 2005). In order to ensure a sufficient supply of these renewable fuels, efforts should be made to evaluate their economic feasibility in states, such as North Dakota, that have vast feedstock resources.

Utilizing North Dakota's canola resources for biodiesel production

The biodiesel industry is already well established in the European Union (E.U.) and is rapidly growing in the United States. The conversion of plant oils and animal fats to biodiesel (methyl ester of long chain fatty acids) is relatively simple and far more efficient than the process for fuel ethanol (another popular renewable fuel). The soybean industry is the chief proponent of biodiesel in the United States because of the current low price of soybean oil, and will likely remain a major player in United States biodiesel production. However, soybeans are not well adapted to the cool, dry climate of western and northern North Dakota—especially when compared to other minor oil crops such as canola (a

variation of rapeseed). Also, canola is a far richer source of oil and yields much more oil per hectare.

The E.U. Commission has called for biofuels to account for 2% of E.U. fuel use in 2005 and 5.75% by 2010 (United States Department of Agriculture, 2003). Most of that increase will likely be biodiesel, due in part to the strong demand for diesel in the E.U. Those goals are considered very ambitious; nevertheless, given the recent growth trend for biodiesel in the E.U., attaining a level of 5% of fuel use within 20 years is quite plausible. Were biodiesel to attain only 2% of United States diesel consumption, an annual production of over 600 million gallons of biodiesel would be required. If 5% of that market is captured by biodiesel from canola and other oilseeds produced on the Northern Plains, total acreage of those crops could be increased by 400,000 acres (Slayton, 2002). This would have a very positive, pronounced impact on the economy of the region. Thus, an evaluation to determine the costs of producing biodiesel from canola oil in North Dakota is necessary. This evaluation can also determine how this region should participate in the United States biodiesel industry.

Therefore, the primary focus of this study is to evaluate the feasibility of biodiesel production in North Dakota with canola oil as the feedstock. North Dakota, with 980,000 acres of canola harvested area in 2005, tops the ranking list of all United States canola-producing states—representing 90% of the total canola production of the United States (ND Agricultural Statistics Services, 2005). With such huge feedstock resources, North Dakota can easily take the United States biodiesel sector to the next level by having a biodiesel production facility. While the current study was being performed, announcements

have already been made to construct two biodiesel production plants in North Dakota, with 30 and 50 MGY capacities and canola oil as the feedstock.

Application of the process modeling and engineering economic approach

Efforts have been made 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 apply 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. 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 processes.

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. Performing the mass and energy balance analysis of the full process is a key step toward obtaining economic cost data for performing economic feasibility studies.

This study applies the process modeling approach together with the application of process economics principles to evaluate the detailed economic feasibility of biodiesel production in North Dakota. The study analyses the biodiesel production process leading to the development of a biodiesel process model using spreadsheets and other software tools. The model incorporates material and energy balances and other physical-chemical principles and also quantifies biodiesel production inputs and outputs. The results from this process model help to determine biodiesel equipment specifications and utility requirements with crude canola oil as the biodiesel feedstock in North Dakota. Equipment specifications are utilized to calculate the detailed equipment costs, which are the basis for calculating the total capital investment required for setting up the biodiesel production plant. The inputs and outputs of the process model and the calculated utilities requirement are combined with the local cost data to determine the total annual biodiesel cost/gallon, thus, evaluating the economic feasibility of the biodiesel production in North Dakota. All these calculations are performed for two production capacities—5 MGY and 30 MGY—to determine the presence of economies of scale in the biodiesel production process.

Thesis organization

This thesis consists of two papers. They are Paper 1: Process Model for Biodiesel Production from Various Feedstocks, and Paper 2: Economic Feasibility of the Canola Biodiesel Production Plant in North Dakota. In the first paper, a biodiesel process model is presented with commonly used spreadsheet software and process-engineering principles. The model is based on a continuous two-stage base-catalyzed biodiesel production process.

In the second paper, the mass flow rates data from the developed biodiesel process model have been linked to cost data for evaluating the economic feasibility of biodiesel production in North Dakota with canola oil as the feedstock. The study was conducted to evaluate the feasibility of two canola biodiesel production plants with 5 and 30 MGY capacities, respectively. Total capital investment estimates are based on the purchased biodiesel equipment cost, which is calculated from the equipment specifications generated from the process model results and the process design principles. Appendix A includes the detailed calculations performed for generating these equipment specifications, and Appendix B includes the cell formulas for all the spreadsheets that were prepared for the biodiesel process model.

References

- Coltrain, D. Biodiesel: Is it worth considering? Unpublished manuscript presented at Risk and Profit Conference, Kansas State University, Manhattan, KS, p. 19, April 15-16, 2002.
- Frazier Barnes & Associates. 2003. Mississippi statewide biodiesel feasibility study report. Prepared for The Mississippi Biomass Council, Memphis, TN.
- Fryer, P. J., Pyle, D. L. & Rielly, C.D. 1997. Chemical Engineering for the Food Industry. New York, NY: Blackie Academic and Professional.
- Knothe, G., Krahel, J. & Gerpen, J. V. 2005. The Biodiesel Handbook. Champaign, IL: AOCS Press.
- Koo-Oshima, S. N., Hahn, J. V. & Van Gerpen, J. 1998. Comprehensive health and environmental effects of biodiesel as an alternative fuel. Iowa State University, Ames, IA.

New ultra low sulfur diesel regulations. 2006. Diesel fuel frequently asked question Chevron corporation. Retrieved July 30, 2006, from <http://www.chevron.com/products/prodserv/fuels/diesel/ulsd.shtml>

North Dakota canola acreage and production analysis. June 2005. North Dakota Agricultural Statistics Service report. Retrieved August 9, 2005, from <http://www.nass.usda.gov/nd/canola74.pdf>

Quick, G. R. 1989. Oilseeds as Energy Crops in Oil Crops of the World. McGraw-Publishing Co. pp. 118-131.

Sheehan, J., Camobreco, V., Duffield, J., Graboski, M. & Shapouri, H. 1998. Life cycle inventory of biodiesel and petroleum diesel for use in an urban bus. NREL/SR-580-24089. Golden, CO: National Renewable Energy Laboratory.

Slayton, A. Choosing the proper biodiesel technology, paper 2198. Proceedings of the 10th Biennial Bioenergy Conference, Boise, ID, September 22-26, 2002.

United States Department of Agriculture – Production Estimates and Crop Assessment Division Foreign Agricultural Service. November 2003. Rapeseed Production May Benefit from New EU Directive report. Retrieved March 7, 2006, from <http://ffas.usda.gov/pecad2/highlights/2003/11/biodieseldirective/index.htm>

Urbanchuk, J.M. 2001. An economic analysis of legislation for a renewable fuels requirement for highway motor fuels. AUS Consultants. Mt. Laurel, New Jersey.

U.S. Department of Energy. Energy pact act of 2005 (EPACT). Retrieved January 17, 2005, from <http://www.energy.gov/taxbreaks.htm>

U.S. Department of Energy/EIA. November 2005. Retrieved May 26th, 2006, from http://tonto.eia.doe.gov/dnav/pet/pet_sum_crdsnd_adc_mbbbl_m.htm

- Van Wechel, T., Gustafson, C. R. & Leistritz, F. L. 2002. Economic feasibility of biodiesel production in North Dakota. Agribusiness and Applied Economics Report No. 505. Fargo, ND: North Dakota State University, Agricultural Experiment Station.
- Zhang, X., Peterson, C.L., Reece, D., Moller, G. & Haws, R. 1995. Biodegradability of biodiesel in the aquatic environment. ASAE Paper No. 956742. St. Joseph, MI: ASAE.
- Ziejewski, M. 1985. The effect of the level of unsaturation and of alcohol modifications of plant oil fuels on the long term performance of a direct injected diesel engine. Ph.D. thesis, North Dakota State University, Fargo, North Dakota, USA.

PAPER 1

PROCESS MODEL FOR BIODIESEL PRODUCTION FROM VARIOUS FEEDSTOCKS

ABSTRACT

Researchers need tools to evaluate and compare different available feedstocks, and process parameters and modifications for biodiesel production. To address this need, this study presents a biodiesel process model that uses common 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 are 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 are 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.

INTRODUCTION

Chemistry of the transesterification reaction

A process known as “transesterification” converts triglyceride oils/fats to alkyl esters to produce biodiesel from vegetable oils or animal fats. Figure 1 shows the basic transesterification reaction.

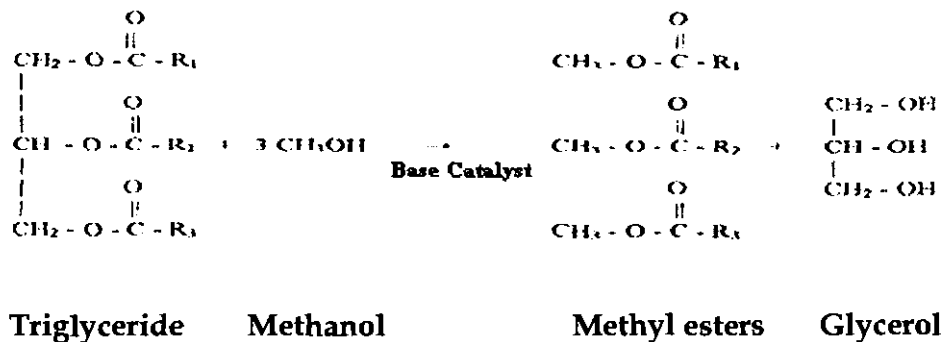


Figure 1. Transesterification reaction.

Alcohol reacts with oil in the transesterification process to release three “ester chains” (alkyl esters) from the glycerin backbone of each triglyceride. The esters are termed “biodiesel” and have better fuel properties—such as less viscosity and more volatility—than the original oil or fat. The reaction requires heat and a catalyst to achieve complete conversion of the oil or fat into the alkyl esters and glycerol. R_1 , R_2 , and R_3 represent the fatty acid chains. Five fatty acids are most common in vegetable oils and animal fats (others are present in small amounts):

Palmitic: $R = -(\text{CH}_2)_{14}-\text{CH}_3$ (16:0)

Stearic: $R = -(\text{CH}_2)_{16}-\text{CH}_3$ (18:0)

Oleic: $R = -(\text{CH}_2)_7\text{CH}=\text{CH}(\text{CH}_2)_7\text{CH}_3$ (18:1)

Linoleic: $R = -(\text{CH}_2)_7\text{CH}=\text{CH}-\text{CH}_2-\text{CH}=\text{CH}(\text{CH}_2)_4\text{CH}_3$ (18:2)

Linolenic: R= $-(CH_2)_7CH=CH-CH_2-CH=CH-CH_2-CH=CH-CH_2-CH_3$ (18:3)

Two numbers separated by a colon designate the fatty acids' properties. The first number designates the number of carbon atoms, and the second number designates the number of double bonds. Table 1 shows the percentages of each fatty acid present in common oils and fats. The glycerol molecule has three carbon atoms, together with five hydrogen atoms and three OH or hydroxyl groups (Van Gerpen et al., 2003).

Table 1. Typical fatty acid composition of various oils and fats (Erickson, 1995)

Oil or Fat	14:0	16:0	18:0	18:1	18:2	18:3
Soybean		11	4	23	54	8
Canola		4.3	1.3	59.9	21.1	13.2
Lard	1-2	28-30	12-18	40-50	7-13	0-1
Yellow Grease	2.43	23.24	12.96	44.32	6.97	0.67

Biodiesel production technology

Preparation of feedstock and other raw materials

Biodiesel can be produced from a wide variety of fat and oil sources such as vegetable oil (soybean, canola, sunflower, safflower, etc), animal fats (beef tallow, lard, poultry fat), restaurant waste oils, and trap grease. These sources have to be prepared before applying the transesterification reaction.

Crude vegetable oil processing

Degumming, refining, bleaching, and deodorization are the steps commonly involved in processing crude oil, and after processing, the oils are known as RBD oils

(refined, bleached, and deodorized). Studies have shown that high-quality biodiesel can be produced from crude oils. However, use of partially or fully refined oil simplifies the biodiesel production process. Therefore, some producers choose to use degummed oils or even RBD oils for production (Sheehan et al., 1998; Van Gerpen et al., 2003).

Degumming/refining

Refining consists of two steps. The first step is degumming, which removes the phospholipids (typically 1.5-2.5% of the crude oil by weight) from the crude oil. Refiners remove these phospholipids primarily because some of the compounds, particularly the calcium and magnesium salts of phosphatidic and lysophosphatidic acids, are strong emulsifiers. If these compounds are still present during the later alkali neutralization step, they will inhibit the separation of the soaps and reduce the yield of neutral oil. Phospholipids will also react with water to form undesirable insoluble sediments. Degumming is also an essential process to meet the phosphorus ASTM specifications in biodiesel by removing the phospholipids. Degumming consists of two steps: water degumming and acid degumming. Water degumming involves mixing the crude oil with soft water in a mixing tank to form gums from the hydratable phospholipids. During the acid degumming step, the resulting mixture is heated to 70°C, followed by the addition of phosphoric acid or citric acid, in order to convert the non-hydratable phospholipids to water-soluble phosphatidic acid. Degummed oil is then separated from the gum-water mix using a centrifuge (Erickson, 1995).

The second phase of refining is neutralization or caustic refining, which removes the free fatty acids (typically 0.3-0.7% of the crude oil by weight) present in the crude oil. An alkali solution, usually sodium hydroxide, is added that reacts with the free fatty acids

to produce soaps. The soaps are insoluble in the oil and easily separated by water washing. The alkali solution also neutralizes any acid remaining from the degumming stage. The byproduct of caustic refining is a mixture of soap, water and oil known as soapstock (Erickson, 1995; Sheehan et al., 1998).

Bleaching

The primary purpose of bleaching is to remove the color pigments from the oil. Bleaching also helps to remove remaining soap, trace metals, phosphatides, and sulfur compounds. Bleaching involves mixing bleaching clays with the oil and agitating for 10-30 minutes (Erickson, 1995).

Deodorization

Deodorization removes the trace components that give an unpleasant taste and odor. Deodorization is essentially a steam distillation process that occurs at high temperatures (210-274°C) and low pressure (1-6 mm Hg) (Erickson, 1995).

Preparation of alcohol and catalyst

Other raw materials required for the transesterification reaction for biodiesel production are alcohol and catalysts.

Alcohol

The most commonly used primary alcohol used in biodiesel production is methanol, although other alcohols, such as ethanol, isopropanol, and butyl, can be used. A key quality factor for the primary alcohol is the water content. Water interferes with transesterification reactions and can result in poor yields and high levels of soap, free fatty acids, and triglycerides in the final fuel. The quality requirements for the alcohol are as follows:

1. Moisture content should be less than 0.08% wt/wt.

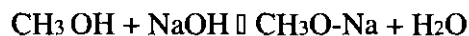
2. Free acids content should be less than 0.0015%.

3. Non-volatiles should be less than 0.003%.

As compared to ethanol, methanol is more commonly used due to its lower boiling point and no azeotrope formation with water. The difference between the boiling points of the two alcohols (64.7°C for methanol and 78.4°C for ethanol) makes the methanol considerably easier to recover than the ethanol at the end of the transesterification reaction. Also, ethanol forms an azeotrope with water, so it is expensive to purify the ethanol during recovery. If the water is not removed it will interfere with the reactions. Methanol recycles easier because it doesn't form an azeotrope. These two factors are the reason that even though methanol is more toxic, it is the preferred alcohol for producing biodiesel (Van Gerpen et al., 2003).

Catalysts

The transesterification reaction requires a catalyst because the alcohol (methanol or ethanol) is sparingly soluble in the oil phase. The catalyst promotes an increase in solubility to allow the reaction to proceed at a reasonable rate. Catalysts may either be base, acid, or enzyme materials. The most commonly used catalyst materials are sodium hydroxide and potassium hydroxide. These base catalysts are highly hygroscopic, and they form chemical water when dissolved in the alcohol reactant based on the following exothermic reaction:



The presence of water (formed in the above reaction) affects the transesterification negatively because it results in the formation of free fatty acids instead of the methyl esters. Because of this reaction, direct use of sodium or potassium alkylate as catalysts is

becoming of greater interest. Acid catalysts are more commonly used for the esterification of free fatty acids. Acid catalysts include sulfuric acid and phosphoric acid (Van Gerpen et al., 2003).

Finding a suitable solid-phase catalyst is a very important current area of research. Such a catalyst would allow the catalyst to be reused, but more importantly, would greatly simplify the refining of the methyl ester (Verkade et al., 2006).

Biodiesel production processes

There are many biodiesel production processes—depending upon the catalyst used, processing technique utilized for the transesterification reaction (batch process or the continuous process) and the feedstock used (low free fatty acid or high free fatty acid). The following are the most commonly used production processes.

Batch processing

Batch processes use a batch, stirred tank reactor for the transesterification reaction. Transesterification requires an alcohol to triglyceride ratio of 3:1 based on the reaction stoichiometry (refer Figure 1). Alcohol to triglyceride ratios from 4:1 to 20:1 (mole:mole) have been reported for batch processes, with a 6:1 ratio most common. This excess alcohol is added to ensure that the transesterification reaction goes to completion. In general, reactions can be encouraged to progress by adding an excess of one of the reactants or by removing one of the products. The reactor may be sealed or equipped with a reflux condenser. The operating temperature is usually close to the boiling point of methanol (around 65°C), although temperatures from 25 °C to 85°C have been reported. The most commonly used catalyst is sodium hydroxide, with potassium hydroxide also being used.

Typical catalyst loadings range from 0.3% to about 1.5% of the total reaction mixture (Van Gerpen et al., 2003).

Thorough mixing is necessary at the beginning of the reaction to bring the oil, catalyst and alcohol into intimate contact. Toward the end of the reaction process, less mixing can help increase the extent of reaction by allowing the inhibitory product, glycerol, to phase separate from the ester – oil phase. Triglyceride conversions of 85% to 94% are reported. Higher temperatures, more time and higher alcohol:oil ratios can enhance the percent completion. Typical reaction times range from 20 minutes to more than 1 hour (Van Gerpen et al., 2003).

Continuous processing

Several processes use intense mixing, with pumps or motionless mixers, to initiate the esterification reaction. A popular variation of the batch process is the use of two continuous stirred-tank reactors in series. The relative volumes of the CSTRs can be varied to allow for a longer residence time in CSTR 1—to achieve a greater reaction. After the initial product glycerol is decanted (glycerol is inhibitory to the transesterification reaction), the reaction in CSTR 2 is rather rapid, with 98% completion not uncommon (Sheehan et al., 1998).

An essential element in the design of a CSTR is sufficient mixing input to ensure that the composition throughout each reactor is essentially constant. This has the effect of increasing the dispersion of the glycerol product in the ester phase. A disadvantage of this mixing is extension of the time required for the phase separation.

Another continuous process occurs in a reactor, referred to as a plug-flow reactor (PFR). Instead of allowing time for the reaction in an agitated tank, the reactor is tubular

and the reaction mixture moves through this reactor in a continuous plug, with little mixing in the axial direction. PFR behaves as if it were a series of many small continuous stirred-tank reactors (CSTRs) chained together. The result is a continuous system that requires lower residence times, as low as 6 to 10 minutes, for completion of the reaction (Van Gerpen et al., 2003.)

High free fatty acid systems

Free fatty acid in feedstock reacts with the catalyst to form soap in a base-catalyzed system. This reaction is tolerable at low free fatty acid levels, but additional processing or the acid catalysis method is required of the feedstocks with high free fatty acid content. The maximum amount of free fatty acids acceptable in a base-catalyzed system is less than 2%, and preferably less than 1%.

Acid-catalyzed processes represent the alternative path of transesterification for the high free fatty acid feedstocks (>2%) such as tallow or yellow grease. Acid catalysts can be used either to directly esterify the free fatty acids or to make esters from soapstock, after first neutralizing the free fatty acids with a base such as sodium hydroxide.

Direct esterification of a high free fatty acid feed requires water removal during the reaction, or the reaction will be quenched prematurely. Also, a very high alcohol to free fatty acid ratio is required—about 40:1. Direct esterification may require rather large amounts of the acid catalyst. The resulting mixture of esters and triglyceride can be used in a conventional base-catalyzed system. Commonly used acid catalysts are sulfuric acid and phosphoric acid.

An alternative approach to utilization of high free fatty acid feedstocks is the use of a base catalyst to deliberately form soap from the free fatty acids. The soap is recovered,

and the oil is dried and used in a conventional base-catalyzed system. This strategy can lead to a false sense of economy. If the soapstock is discarded, the effective price of the feedstock is increased in inverse proportion to the percentage of remaining oil. The soapstock can, however, be converted into esters by using an acid catalyzed reaction. The problem with this strategy is that the soapstock system contains a large amount of water that must be removed before the product esters can meet the biodiesel standard (Van Gerpen et al., 2003).

Treatment and recovery of side streams

Three non-ester side streams must be treated as a part of the overall biodiesel process. These streams are the excess alcohol that is recycled within the process, the glycerol co-product and the wastewater stream from the process.

Methanol recycle

Methanol recycle is necessary because an excess of methanol is required for an effective transesterification reaction system, and failure to recycle the unreacted methanol would be prohibitively expensive. The recovery of the unused methanol also essentially eliminates the emissions of methanol to the surroundings. The emission reduction is needed because methanol is highly flammable and toxic.

In addition, if the methanol is allowed to remain in the system during phase separation, it can act as a phase stabilizer, retarding the rate of gravity separation. It is advantageous to remove the methanol before phase separation. Methanol can be recovered using distillation, either conventional or vacuum, or partially recovered in a single-stage flash.

An alternative to distillation is a falling-film evaporator. Residual methanol in the ester phase can be removed in the water wash step in ester post-processing. Product esters are typically washed with warm (60 °C), softened water to remove soaps and residual methanol (Sheehan et al., 1998; Van Gerpen et al., 2003).

Glycerol refining

Glycerol is recovered and partially refined as a co-product from biodiesel production. The recovered glycerol from the transesterification reaction contains residual alcohol, catalyst residue, carry-over fat/oil, and some esters. Chemical and physical methods are employed to refine the glycerol.

Several factors are important in the chemical refining of glycerol. First, the catalyst tends to concentrate in the glycerol phase where it must be neutralized. The neutralization step leads to the precipitation of salts. Also, the soaps produced in the esterification must be removed by coagulation and precipitation with aluminum sulfate or ferric chloride.

Physical refining removes fatty, insoluble, or precipitated solids by filtration and/or centrifugation and the removal of water by evaporation (Van Gerpen et al., 2003).

OBJECTIVES

The objectives of this study are to collect detailed biodiesel process information and to develop a biodiesel process model based on that information. The model quantifies all the mass flows involved in the biodiesel production process by using the mass balance principles and computer spreadsheets and other software tools. The model also incorporates energy balance principles where necessary to complete the associated mass balances. Model development includes specifying user inputs, identifying the necessary assumptions required for mass and energy balance equations, and performing the detailed underlying calculations. Thus, the various biodiesel process inputs and outputs can be readily quantified at the user-defined production levels.

A base-catalyzed process with two continuous stirred tank reactors (CSTR) for the transesterification reaction was selected for developing the process model, based on a comprehensive literature review (Brown, 2003; Canakci et al., 2001; Erickson, 1995; Knothe et al., 2005; Peterson et al., 1997; Sheehan et al., 1998; Van Gerpen et al., 2003). As the biodiesel industry grows, the new processes will likely be continuous. The model was developed with crude vegetable oils as biodiesel feedstock. The model may be adapted to other feedstocks, such as waste oils and animal fats, but this adaptation is outside the scope of the present article.

MATERIALS AND METHODS

Biodiesel production process description

The selected two-stage continuous biodiesel production process is shown in Figure 2. To facilitate understanding, the process is divided into three main sections: crude oil degumming and refining (Erickson, 1995; Hamm et al., 1999; Gunstone, 2002), transesterification reaction and ester washing (Freedman et al., 1984; Freedman et al., Pryde, 1996; Nouredдини et al., 1997), and methanol recovery and glycerol refining (Sheehan et al., 1998; Singh et al., 2004; Van Gerpen et al., 2003).

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 Figure 2) is heated to 70°C in the heater (A in Figure 2). 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,

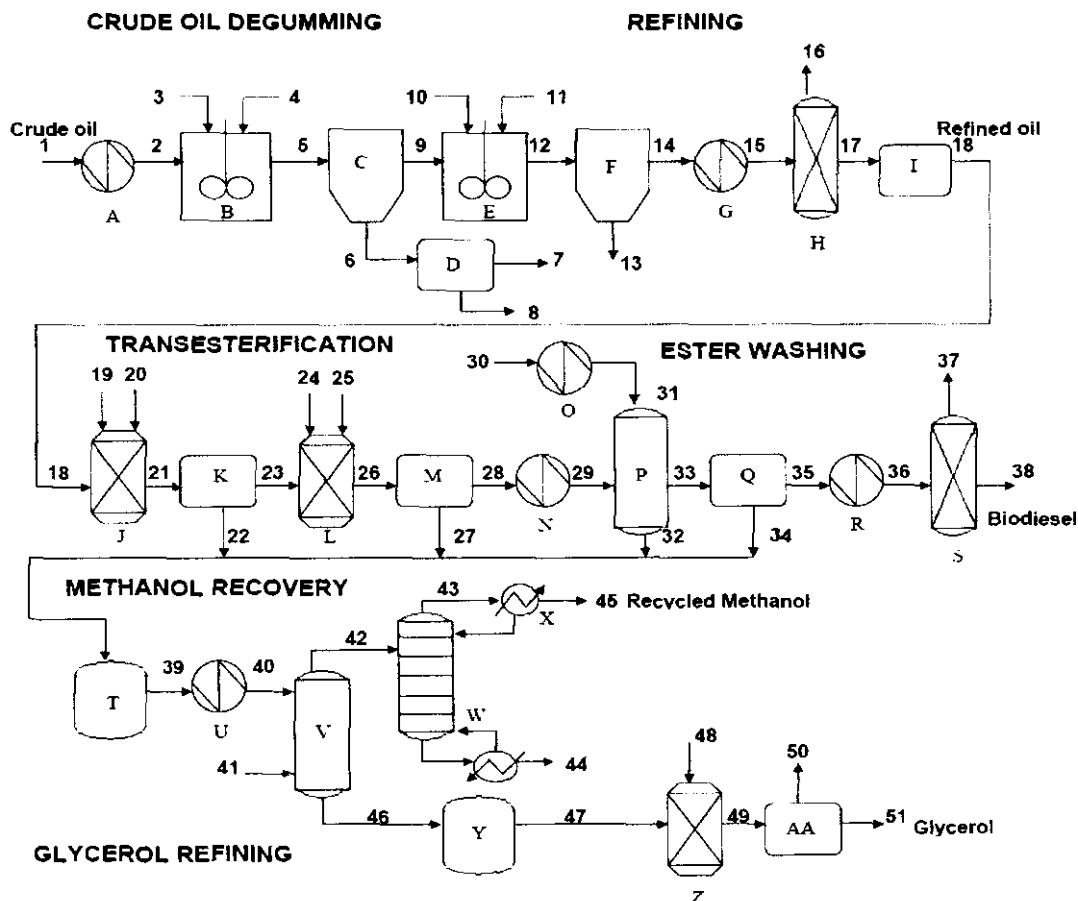


Figure 2. Continuous biodiesel production process used for model development. Equipment: (A) heater, (B) mixing tank, (C) centrifuge, (D) gums/water separator, (E) refining tank, (F) centrifuge, (G) heater, (H) vacuum oil dryer, (I) surge tank, (J) continuous stirred tank reactor (CSTR) 1, (K) decanter 1, (L) CSTR 2, (M) decanter 2, (N) heater, (O) heater, (P) wash columns, (Q) settler tank, (R) heater, (S) vacuum ester dryer, (T) collecting tank, (U) heater, (V) glycerol-alcohol stripper, (W) distillation column/reboiler, (X) reflux condenser, (Y) glycerol hold tank, (Z) acidulation reactor, and (AA) decanter. Streams: (1) crude oil, (2) heated crude oil, (3) phosphoric acid, (4) soft water, (5) mixing tank outstream, (6) gums-water mix, (7) gums, (8) water, (9) degummed oil, (10) NaOH solution, (11) wash water, (12) refining tank outstream, (13) soapstock, (14) centrifuge outstream, (15) heater outstream, (16) water vapor, (17) hot oil, (18) refined oil, (19) sodium methoxide, (20) methanol, (21) CSTR 1 outstream, (22) glycerol phase, (23) ester phase, (24) sodium methoxide, (25) methanol, (26) CSTR 2 outstream, (27) glycerol phase, (28) ester phase, (29) heater outstream, (30) soft water, (31) heated wash water, (32) waste stream, (33) washed esters, (34) aqueous phase, (35) esters, (36) heated esters, (37) water vapor, (38) biodiesel, (39) glycerol/aqueous phase, (40) heater outstream, (41) super heated steam, (42) saturated methanol vapors and saturated steam, (43) methanol vapor, (44) distillation column bottoms, (45) recycled methanol, (46) hot glycerol solution, (47) glycerine, (48) HCl solution, (49) acidulation reactor outstream, (50) waste, and (51) product glycerol.

to remove the remaining water in the oil. The dried, degummed and refined oil is then sent to a surge tank.

Conventional edible oil refining uses above-mentioned steps. The bleaching and deodorizing units are 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 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 (Figure 3):

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

A	B	C	D	E	F	G	H
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
11 Descr.	Crude Oil	Heater Outstream	Phosphoric acid solution	Soft water	Mixing tank Outstream	Gums-water mix	Gums
13							
14 Component							
16 Water (w)			0.02	1.50	1.52	1.51	
17 Methanol (m)							
18 Sodium Methoxide (sm)							
19 Crude Oil							
20 TG (t)	96.00	96.00			96.00	0.48	0.48
21 FFA (f)	0.80	0.80			0.80		
22 Phosphatides (p)	2.00	2.00			0.00		
23 Others (Unsaponifiable matter) (o)	1.20	1.20			1.20		
24 Glycerol (g)							
25 Soaps (s)							
26 Methyl Ester (me)							
27 NaOH (n)							
28 Phosphoric acid (pa)			0.09				
29 HCl (h)							
30 Gums (gu)					2.09	2.09	2.09
31 Steam (st)							
32 NaCl (na)							
33 Total (t)	100.00	100.00	0.10	1.50	101.60	4.07	2.57
43 Fatty acid	C 16	C 16:1	C 18	C 18:1	C 18:2	C 18:3	C 20
44 Triglyceride Molecular weight (mw)	887.34	881.29	891.50	885.45	879.41	873.36	975.68
45 Input Composition of the oil (% age)	4.00	0.50	2.00	60.00	20.00	10.50	1.50
48 Molecular weight fraction	32.29	4.01	17.83	531.27	175.88	91.70	14.64

Figure 3. Screen view of the first eight streams in the process model spreadsheet.

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

Vacuum oil dryer (H): Complete moisture removal takes place from stream 14.

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. A 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 et al., 1999).

Stream 42 contains 100% recovery of the saturated methanol vapors and saturated steam, which flows to the 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 when and where chemical reactions occur.

Mass balance equations are applied to each individual unit for quantifying the components in each stream in the whole production process. It is assumed that no net depletion or accumulation occurs and that all mass is accounted for in the flows shown in Figure 2. Sample calculations follow for three of the 27 units in Figure 2: 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 are 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 presented in the section titled "Crude oil degumming and refining" and are calculated as follows:

Water (*w*) balances are $M_{6w} = 0.995 * M_{5w}$ and $M_{9w} = M_{5w} - M_{6w}$.

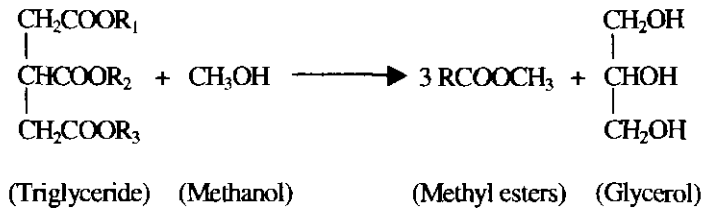
Triglyceride (*t*) balances are $M_{6t} = 0.005 * M_{5t}$ and $M_{9t} = M_{5t} - M_{6t}$.

Gums (*gu*), free fatty acids (*f*), phosphatides (*p*), and other unsaponifiable matter (*o*) balances are $M_{5gu} = M_{6gu}$ and $M_{9gu} = 0$, $M_{5f} = M_{9f}$ and $M_{6f} = 0$, $M_{5p} = M_{9p}$ and $M_{6p} = 0$, and $M_{5o} = M_{9o}$ and $M_{6o} = 0$, 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: $M_6 = M_{6w} + M_{6t} + M_{6gu}$.

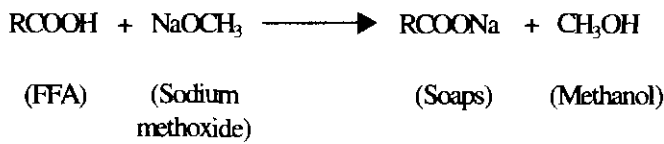
Continuous stirred tank reactor (CSTR) 1

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 other unsaponifiable matter (*o*). The mass flow rates of these components in streams 18, 19, and 20 were calculated earlier in the model. Process information presented in the section titled "Transesterification and Ester Washing" is 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*):

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

Methyl esters (*me*):

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

Glycerol (*g*):

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

Free fatty acids (*f*):

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

Methanol (*m*):

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

Sodium methoxide (*sm*):

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

Soaps (*s*):

$M_{21s} = [(MW_s)/(MW_f)] * (1 * M_{18f})$ (Eq 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} are calculated as weighted averages from the vegetable oil free fatty acid composition data:

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

$$MW_f = \sum_k [MW_k * WF_k] \quad (\text{Eq 9})$$

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} is calculated from the transesterification stoichiometry:

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

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 in the section titled "Methanol Recovery and Glycerol Refining"):

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

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

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

M_{40} and all M_{40j} are known. Component balances include: $M_{42m} = M_{40m}$, $M_{42w} = M_{40w} + M_{41w}$, $M_{42} = M_{42m} + M_{42w}$, $M_{46g} = M_{40g}$, $M_{46sm} = M_{40sm}$, $M_{46t} = M_{40t}$, $M_{46o} = M_{40o}$, and $M_{46s} = M_{40s}$.

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

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

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

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

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

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

where M_i is the mass flow rate (kg/h) of any stream i ; \bar{C}_{p40} is the average specific 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 point; and H_{42s} , H_{42w} , and H_{42m}

are the enthalpies of the components (saturated steam, water vapor, and methanol vapor, respectively) in stream 42. Enthalpies are calculated with 0°C as the reference. In the case of water and methanol vapor, saturated liquid at 0°C is the reference. H_{41} and H_{42w} are determined from superheated steam and saturated steam tables, respectively.

DISCUSSION

Figure 3 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 3 only shows 7 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 2 by choosing a 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 2 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 2 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 primarily with the mass balance calculations, and enthalpy balance calculations in the case of the glycerol-alcohol stripper (V), that are necessary for characterizing all mass flows. A separate spreadsheet is included in the model that addresses 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.

Table 2. 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) ^[a]	Canola Oil (kg/h) ^[b]
Process Inputs		
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
Process Outputs		
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

^a Composition: 96.0% triglycerides, 0.5% free fatty acids, 2.0% phosphatides, and 1.5% other unsaponifiable matter . Fatty acid profile: 11% C16, 4% C18, 23% C18:1, 54% C18:2, 8% C18:3 (Erickson, 1995).

^b Composition: 97.25% triglycerides, 0.5% free fatty acids, 1.25% phosphatides, and 1.0% other 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 et al., 1999).

CONCLUSION

The model described here was developed for analyzing biodiesel production. The model is 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.

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 (for instance, 0.015 million m³ 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.

REFERENCES

- Brown, R. C. 2003. Biorenewable Resources. Ames, IA: Iowa State University Press.
- Canakci, M., & Van Gerpen, J. 2001. A pilot plant to produce biodiesel from high free fatty acid feedstocks. ASAE Paper No. 016049. St Joseph, MI: ASAE.
- Erickson, D. R. 1995. Practical Handbook of Soybean Processing and Utilization. Champaign, IL: AOCS Press.
- Freedman, B., Butterfield, R. O. & Pryde, E. H. 1996. Transesterification kinetics of soybean oil. J. American Oil Chemists Society 63(10): 1375-1380.
- Freedman, B., Pryde, E. H. & Mounts, T. L. 1984. Variables affecting the yields of fatty esters from transesterified vegetable oils. J. American Oil Chemists Society 61(10): 1638-1643.
- Fryer, P. J., Pyle, D. L. & Rielly, C. D. 1997. Chemical Engineering for the Food Industry. New York, NY: Blackie Academic and Professional.
- Gunstone, F. D. 2002. Vegetable Oils in Food Technology: Composition, Properties and Uses. Oxford, U.K.: Blackwell Publishing.
- Hamm, W., & Hamilton, R. J. 1999. Edible Oil Processing. Sheffield, U.K.: CRC Press.
- Hernandez, E., and Lusas, E.W. 1996. Practical short course in Processing of Vegetable Oils. College station, TX: The Texas A&M University System.
- Knothe, G., Krah, J. & Gerpen, J. V. 2005. The Biodiesel Handbook. Champaign, IL: AOCS Press.

- Noureddini, H., & Zhu, D. 1997. Kinetics of transesterification of soybean oil. *J. American Oil Chemists Society* 74(11): 1457-1463.
- Peterson, C.L., Reece, D.L., Hammond, B.L., Thompson, J. & Beck, S.M. 1997. Processing, characterization, and performance of eight fuels from lipids. *Applied Engineering in Agriculture* 13(1): 71-79.
- Sheehan, J., Camobreco, V. Duffield, J., Graboski, M. & Shapouri, H. 1998. Life cycle inventory of biodiesel and petroleum diesel for use in an urban bus. NREL/SR-580-24089. Golden, CO: National Renewable Energy Laboratory.
- Singh, A. P., Thompson, J. C. & He, B. 2004. A continuous-flow reactive distillation reactor for biodiesel preparation from seed oils. ASAE Paper No. 046071. St Joseph, MI: ASAE.
- Van Gerpen, J., Clemens, D., Knothe, G., Shanks, B. & Pruszko, R. 2003. Biodiesel production technology. Ames, IA: Iowa State University, Biodiesel Workshop.
- Van Gerpen, J.H. & Dvorak, B. The effects of phosphorus level on the total glycerol and reaction yield of biodiesel, presented at Bioenergy 2002, The 10th Biennial Bioenergy Conference, Boise, ID, Sept. 22-26, 2002.
- Verkade, J.G., Shanks, B.H. & Y. Lin, V.S. 2006. Nanoporous solid catalysts for conversion of soybean oil to biodiesel. Undergoing project, Department of Chemistry, Iowa State University. Ames, Iowa.

PAPER 2

ECONOMIC FEASIBILITY OF A CANOLA BIODIESEL PRODUCTION PLANT IN NORTH DAKOTA

ABSTRACT

The economic feasibility of biodiesel production in North Dakota with canola oil as the feedstock is assessed. Estimations for capital investment cost and total annual biodiesel product cost are conducted for two canola biodiesel production plants with 5 million gallons per year (MGY) and 30 MGY capacities. These capacities were selected based on North Dakota's and its neighboring states' biodiesel demands, respectively. Capital investment cost analysis shows the presence of considerable economies of scale for the biodiesel production process for the two capacities. These cost calculations are based on the purchased equipment cost that is calculated from equipment specifications. Total annual biodiesel product cost analysis shows that the major portion (>80%) of the total product cost is the raw materials cost similar to the analysis of previous economic feasibility studies. Cost benefits from the economies of scale are still present for the fixed charges, general expenses and the manufacturing costs (other than the raw material costs) in the annual product cost calculations. Finally, based on the gross profit evaluation for both the plants, it is concluded that it is more worthwhile to invest in the 30 MGY production plant because of the greater cost returns from the economies of scale benefits. The results are more encouraging after the incorporation of the federal biodiesel tax incentive and favor the investment for biodiesel production in North Dakota.

INTRODUCTION

Currently, a dozen companies have made investments in U.S. biodiesel production plants, and eight economic feasibility studies from different states have been published to foster United States biodiesel growth. A primary drawback to further expansion is high production cost (nearly 75% of which is constituted by the feedstock cost) which, in turn, results in high biodiesel prices. Fortenbery (2005) recently assessed biodiesel potential in Wisconsin and concluded that biodiesel does not compete with petroleum diesel on a price per gallon basis and that the investment in a biodiesel plant is essentially a speculative bet on favorable public policy initiatives occurring in the near future. All the previous economic feasibility study groups have relied on a "black box" approach in which the reader cannot clearly define the underlying calculations performed for evaluating biodiesel economic feasibility. The studies have provided only the overall input and output data of biodiesel production collected from private consultancy firms (Frazier Barnes & Associates, 2003; VanWechel et al., 2002), and they have utilized those data for performing the studies. To resolve the above issues and in an effort to provide a complete understanding of the biodiesel production process and the method of evaluating the biodiesel economic feasibility study, the present study has been undertaken. This study utilizes the developed biodiesel process model (Tapasvi et al., 2005) and the principles of engineering economics for evaluating the economic feasibility of biodiesel production in North Dakota with canola oil as the feedstock.

Biodiesel legislation

Federal biodiesel legislation

On August 8, 2005, President Bush signed the Energy Policy Act (EPAAct) of 2005 with the biodiesel credit. This credit benefits small agri-biodiesel producers by giving them a 10-cent per gallon tax credit for up to 15 MGY of agri-biodiesel produced. The credit is effective until the end of 2008 (US Department of Energy, 2005). Previously, on April 25, 2002, the U.S. Senate approved the Energy Bill, S. 517, which includes provisions for biodiesel. These provisions contain a biodiesel excise tax incentive, renewable fuels standard, blenders tax credit, amendment of the EPAAct and federal fleet use requirements (S.517, H.R.4, *Biodiesel Bulletin*, May 1, 2002). The Biodiesel Excise Tax Incentive provides blenders of biodiesel with a 1-cent reduction in diesel excise tax for every percentage of biodiesel made from virgin vegetable oil, up to a 20% content. A Blender's Tax Credit also offers a 1/2-cent per percent up to a 20% tax credit for biodiesel made from recycled oils and animal fats. The Renewable Fuels Standard specifies biodiesel as an eligible fuel that can help meet 5 billion gallons per year fuels standard. The EPAAct amendment removes the 50% limit on biodiesel use for government fleets. Finally, the legislation requires the federal government to use biodiesel when it is cost competitive (*Biodiesel Bulletin*, May 1, 2002).

State biodiesel legislation

Several important legislative measures designed to encourage the use of biodiesel in individual states have been introduced.

Minnesota

On March 15, 2002, the state of Minnesota passed legislation creating a new law requiring a majority of the state's diesel to include 2% soy biodiesel beginning June 30 of 2005. Minnesota is the first state to require the use of biodiesel (*Biodiesel Bulletin*, March, 29, 2002). On September 29, 2005, the state successfully implemented this initiative, which blends 2% biodiesel (B2) throughout its entire diesel fuel supply. Minnesota has exceeded the legislative requirement that the state have biodiesel production capacity of at least eight MGY. The state now leads the nation with a biodiesel production capacity of 63 MGY (National Biodiesel Board).

On December 23, 2005, in response to a request from the Minnesota Trucking Association and the Minnesota Biodiesel Council, the Minnesota Department of Commerce issued a 21-day suspension of the state's biodiesel fuel mandate. The move came after receiving some reports showing that soy-based biodiesel was clogging fuel filters, especially in extremely cold conditions. Excessive glycerin in few soy biodiesel lots was identified as the root cause of the problem. The suspension eventually ended on January 13, 2006, after the replacement of low-quality fuel lots with fresh biodiesel supplies (Soygrowers news room, January, 2006).

North Dakota

On April 22, 2005 Gov. John Hoeven signed a comprehensive package of legislation into law designed to accelerate production of biodiesel (North Dakota Office of Governor, press release, April 22, 2005):

1. A \$1.2 million Biodiesel PACE program, which will provide up to \$650,000 in interest buy-down for a biodiesel plant, by providing \$400,000 with \$250,000 in regular PACE.
2. An income tax credit of 10% per year for up to 5 years, or 50% of direct costs, for biodiesel sales equipment such as pumps, hoses and tanks purchased by retailers.
3. An income tax credit for any fuel supplier that blends biodiesel fuel to the ratio of 5%, or B5. The credit equals five cents per gallon of the blended fuel B5.
4. A sales tax exemption on the sale of new equipment to any facility that will enable the facility to sell diesel fuel containing at least 2% biodiesel fuel.

Other states

Although the Minnesota requirement stands out as one of the first significant pieces of public policy, several other states have also passed significant legislation to encourage biodiesel supply and demand (Minnesota mandate press release NBB news, 2005):

1. Arkansas: Passed a fund granting up to 10-cents per gallon of biodiesel produced.
2. Hawaii: Lowered state excise tax for biodiesel blends.
3. Illinois: Enacted a partial state sales tax exemption for biodiesel blends from B1-B10 and a full exemption for B10-B100, through 2013.
4. Indiana: Provides an expansion of state tax credits for biodiesel producers, blenders, and retailers.
5. Missouri: Determined that qualified biodiesel producers are eligible for a monthly grant of 30-cent per gallon of biodiesel for the first 15 million gallons

produced annually, or 10-cent per gallon of biodiesel up to 15 million gallons produced beyond their initial 15 million gallons.

6. Pennsylvania: Developed an Alternative Fuels Incentive Fund to provide grants to schools, municipalities, political subdivisions, non-profits, LLCs and partnerships for purposes including incremental purchase costs of B100 and B20, refueling infrastructure and vehicle retrofitting. The state also reimburses qualified renewable fuels producers up to 5-cents per gallon of the blended fuel.
7. Texas: Provides a production incentive grant of a net 16.8 cents per gallon of biodiesel produced for 18 MGY per plant. Also exempts biodiesel portion of biodiesel blends from state excise tax.

Proposed North Dakota biodiesel production plants

While the current study was being performed, announcements have already been made to construct two biodiesel production plants in North Dakota with canola oil as the feedstock.

1. On March 22, 2005, Senator Kent Conrad announced that a private company, North Dakota Biodiesel Inc., will construct a biodiesel production plant in Minot—an investment that will utilize North Dakota's natural resources, pump millions of dollars into the local economy and create several jobs. The facility will be able to produce around 30 MGY of premium biodiesel annually from more than 355,000 acres of North Dakota-grown canola (Conrad, 2005).
2. On October 5, 2005, Gov. John Hoeven announced that the Archer Daniels Midland Company will construct a 50 MGY biodiesel plant adjacent to the company's canola crushing plant in Velva, ND. The new facility will convert

canola oil from the plant into biodiesel fuel, and use the equivalent of 600,000 acres or 400,000 – 450,000 tons of canola, adding further value to the crop. The facility has tripled production of canola oil in recent years. The plant would create 12 new jobs (North Dakota State Government, 2005). The planned capacity was later increased to 85 MGY.

LITERATURE REVIEW

Biodiesel market potential of North Dakota and neighboring states

Three immediate opportunities for biodiesel market growth in North Dakota are 1) agriculture, 2) construction equipment/industrial, and 3) state fleet. Agriculture represents the largest market segment opportunity for biodiesel in North Dakota. Increasing numbers of farmers are becoming familiar with its technical properties and merits. The cold flow limitation of biodiesel is of minor consequence to farmers as the majority of their field operations are performed during the warm season. North Dakota farmers also have a strong commitment to adding value to their crops and strengthening rural economic development. To estimate total biodiesel market potential, this study assumes a majority of the diesel fuel sold for uses in agriculture, construction, and the state fleet contains 2% biodiesel. (VanWechel et al., 2002).

Table 3 summarizes the petro-diesel utilization for North Dakota during the last 7 years. Petro-diesel usage in North Dakota is nearly constant over the years, averaging 400 MGY. The agriculture, construction and state fleet segment of the diesel usage is roughly 40% of the total usage (VanWechel et al., 2002), and therefore, equals 160 MGY. Assuming 2% biodiesel blends (B2), the annual biodiesel usage for North Dakota equals 3.2 MGY. After including 50% excess capacity for future market demands, a 5 MGY biodiesel production plant will be appropriate in North Dakota.

If neighboring states (in the Northern Plains), such as Montana, South Dakota, Manitoba, Saskatchewan and Minnesota, are included as potential biodiesel consumers for the biodiesel produced in North Dakota, a 30 MGY biodiesel production plant will be

considered appropriate, again after assuming a 2% blend of biodiesel. Larger process plants typically deliver a less expensive end product due to the economies of scale benefits.

Table 3. North Dakota petro-diesel utilization (Galster, 2004)

Fiscal Year	Total Taxable Gallons
July 1997 – June	415,182,790
July 1998 – June 1999	395,195,176
July 1999 – June 2000	373,268,102
July 2000 – June 2001	387,150,753
July 2001 – June 2002	401,303,264
July 2002 – June 2003	378,812,660
July 2003 – June 2004	411,844,012

Assessing the canola oil availability and supply

Most of the canola production in the United States is concentrated in North Dakota with a total production of 1.2 billion pounds in 2004. The oil content of the canola crop is around 40%, and the oil extraction efficiency is greater than 90% (Mag, 2005). Using these values and assuming 98% transesterification efficiency for biodiesel production, the North Dakota canola resources will easily support the 5 MGY biodiesel production plant. Figure 4 shows that the North Dakota canola production has been quite consistent in recent years, with production acres always more than 800,000 since 1998 (North Dakota Agricultural Statistics Service, 2005).

Neighboring regions, such as Manitoba and Saskatchewan, also produce much canola, with 2005 annual production of 1261.0 and 4633.4 thousand tons, respectively (Canola Council of Canada, 2005). These canola resources can easily support a 30 MGY biodiesel production plant in North Dakota after transporting the canola oil feedstock from these states. Thus, it can be concluded that sufficient canola feedstock resources are available for both production plant capacities—5 MGY and 30 MGY—and that the current economic feasibility study is not limited by the feedstock availability constraints.

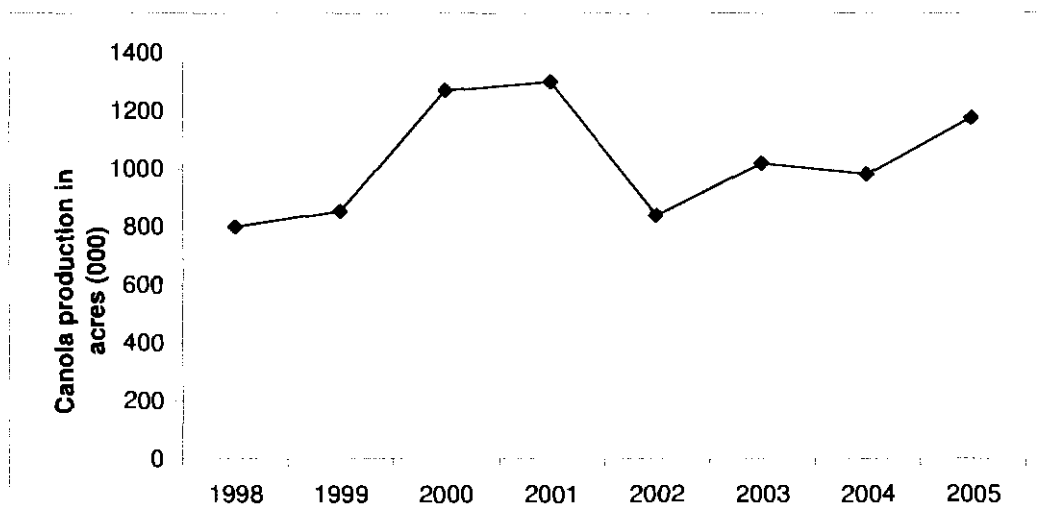


Figure 4. North Dakota canola production trends: 1998-2005 (National Agricultural Statistics Service, 2005).

Review of previous economic feasibility studies

Economic feasibility studies published about biodiesel production in different parts of United States are summarized in Table 4. The method adopted for performing these studies was almost the same and included the following steps:

1. Evaluating feedstock availability

Within a particular region, research was conducted to ensure sufficient feedstock availability for the biodiesel production. The data, such as soybean or canola crop production statistics, was obtained from the United States Department of Agriculture and Census Bureau sources.

2. Process selection

The operating strategy involves the selection of the operation system to be used for the biodiesel production. Operating systems include either a batch system or a continuous system. The catalyst to be used is also decided, i.e. whether to use a base catalyst or an acid catalyst for the transesterification reaction.

The choice of batch and continuous systems is commonly determined by the amount of feedstock to be utilized for biodiesel production. Batch reactions are favored for smaller capacity plants (< 5 MGY) that may operate only one or two shifts per day and the continuous reactions are favored for larger capacity plants

Table 4. Previous economic feasibility studies (* MGY = million gallons per year)

	Study-1	Study-2	Study-3	Study-4	Study-5
Authors	Van Dyne et al.(1992)	Van Wechel et al. (2002)	Fortenbery (2005)	English et al. (2002)	Shumaker et al.(2000)
Region	Southeastern U.S.	North Dakota	Wisconsin	Tennessee	Georgia
Oil used	Canola	Soybean	Soybean	Soybean	Soybean
Capacity	1 MGY	5 MGY	10 MGY	13 MGY	15 MGY

where 24 hour, 7 days a week operation reduces the overall size of the operating equipment.

Base catalysts are used when the feedstock has less than 1% free fatty acid content, e.g. soybean and canola oils and acid catalysts are used when feedstock has greater free fatty acid content e.g. animal fat for the transesterification reaction.

As all these studies use soybean or canola oil as feedstock, base-catalyzed processes with either sodium hydroxide or potassium hydroxide as the base and, except Study 1 performed by Van Dyne which employed a batch operating system due to small capacity, all the other studies consider the continuous operating system for the biodiesel production.

3. Quantification of process inputs and outputs

This step includes the analysis of different biodiesel production process inputs and outputs. The process inputs are feedstock, alcohol, chemicals, catalyst and utilities—such as power source, electricity and water. The process outputs include alkyl esters (biodiesel), glycerol and the other by-products such as soapstock and wash water. Application of process engineering concepts, such as material and energy balances, can be used to calculate relative quantities of inputs and outputs. None of the studies systematically employed engineering principles for these calculations, and the data was collected only from local biodiesel/oilseed processing companies and some consultancy firms.

4. Total capital investment and biodiesel production cost analysis

Data from Step 3 combined with the data from biodiesel manufacturers and

suppliers is used to calculate the total capital investment, as well as the production costs, for a biodiesel plant. The data has been collected from biodiesel producers and consultancy firms that cannot be considered entirely reliable, as these sources might have personal interests behind providing that data, and production details are considered proprietary information that most biodiesel producers are unwilling to share due to fear of competition (VanWechel et al., 2002). Therefore, large variations are seen in the production costs reported. Thus, the economic feasibility studies cannot be regarded as an absolute for the complete biodiesel production cost analysis.

Nearly all the studies have concluded that the main challenge for biodiesel growth is the high and uncertain raw material costs. This can be seen from the production cost analysis in Figure 5, which shows that in all the studies the raw material or the feedstock cost is equal to or more than 75% of the total production costs.

Table 5 provides a summary of the results from these studies. It shows the raw material costs/lb, assumed capacity of the production plant in MGY and the final biodiesel (B100) price calculated per gallon in these studies. It can be seen that generally as the raw material cost decreases and the capacity of the production plant increases, the final biodiesel price decreases.

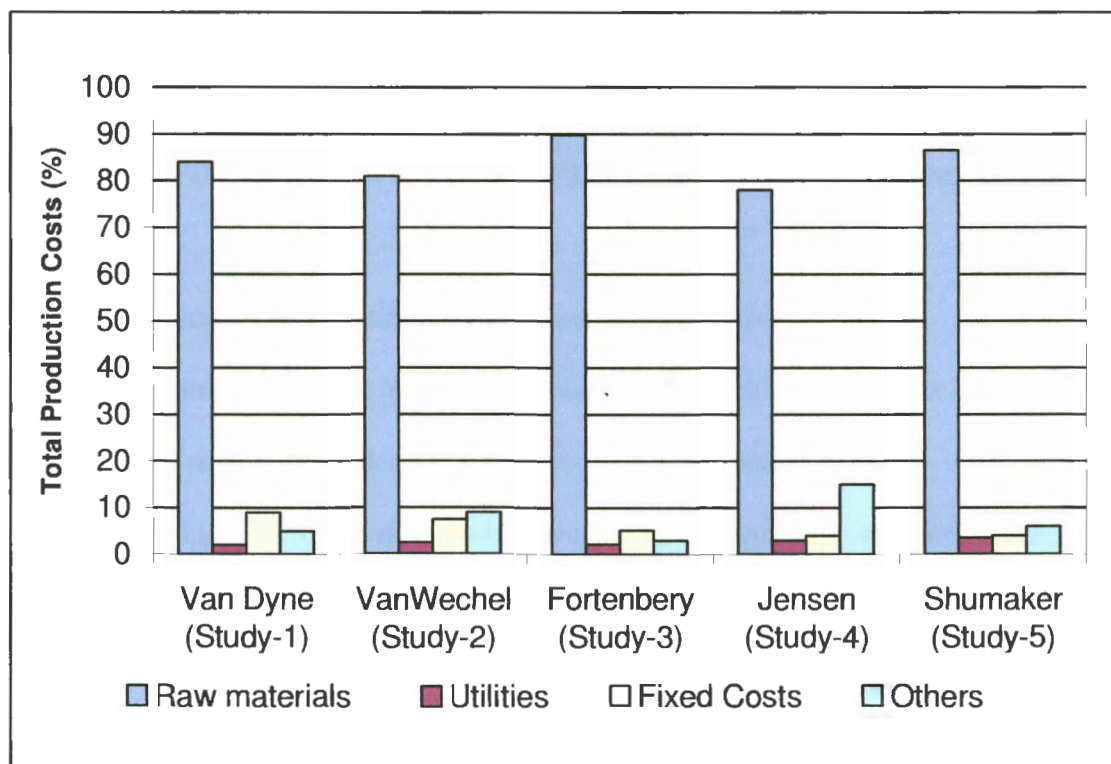


Figure 5. Production cost analysis from economic feasibility studies.

Table 5. Results of the previous economic feasibility studies

	Crude Vegetable Oil (\$/lb)	Capacity (MGY)	Biodiesel Price (\$/Gallon)
Van Dyne Study -1	0.22	1	2.19 - 3.96
VanWechel Study -2	0.17 – 0.25	5	2.02 - 2.64
Fortenbery Study -3	0.10-0.25	10	0.98 - 2.18
Jensen Study -4	0.14	13	1.53
Shumaker Study-5	0.10 – 0.25	15	1.11 – 2.21

OBJECTIVES

To evaluate the economic feasibility of a canola biodiesel production plant in North Dakota with two plant sizes (5 and 30 MGY), the present study has the following two objectives:

1. Estimation of capital investment cost for both plant capacities.
2. Estimation of total biodiesel product cost for both plant capacities.

These estimates are based on the systematic approach of process engineering and economics principles to provide readers with a transparent approach for evaluating the economic feasibility of biodiesel production in North Dakota.

MATERIALS AND METHODS

This economic feasibility study utilizes all the biodiesel production details, flowsheets, and technical assessments as selected by Tapasvi et al. (2005). All the cost calculations were performed using the transparent and systematic approach of process engineering and economic principles (Peters et al., 2003). The operating hours for both plant capacities have been assumed to be 8000 hours/year (Zhang et al., 2003).

Capital investment cost estimate

Capital needed to create the manufacturing and plant facilities is called the fixed-capital investment (FCI). Capital necessary for the operation of the plant is termed working capital (WC). The sum of the fixed-capital investment and the working capital is the total capital investment (TCI). Fixed-capital investment is further subdivided into a manufacturing fixed-capital investment, also known as direct cost, and non-manufacturing fixed-capital investment, also known as indirect cost. Figure 6 explains these costs in detail.

Direct cost calculations

Direct costs include the purchased equipment costs, equipment installation costs, instrumentation and control costs, piping costs, electrical systems costs, buildings costs, and the service facility costs. The cost of purchased equipment is the basis for calculating all the direct costs involved in a biodiesel production facility. This is referred to as the “study estimate,” or the “factored estimate,” as it is based on the knowledge of major items of equipment involved in the production process with a probable estimate accuracy of up to $\pm 30\%$ (Peters et al., 2003).

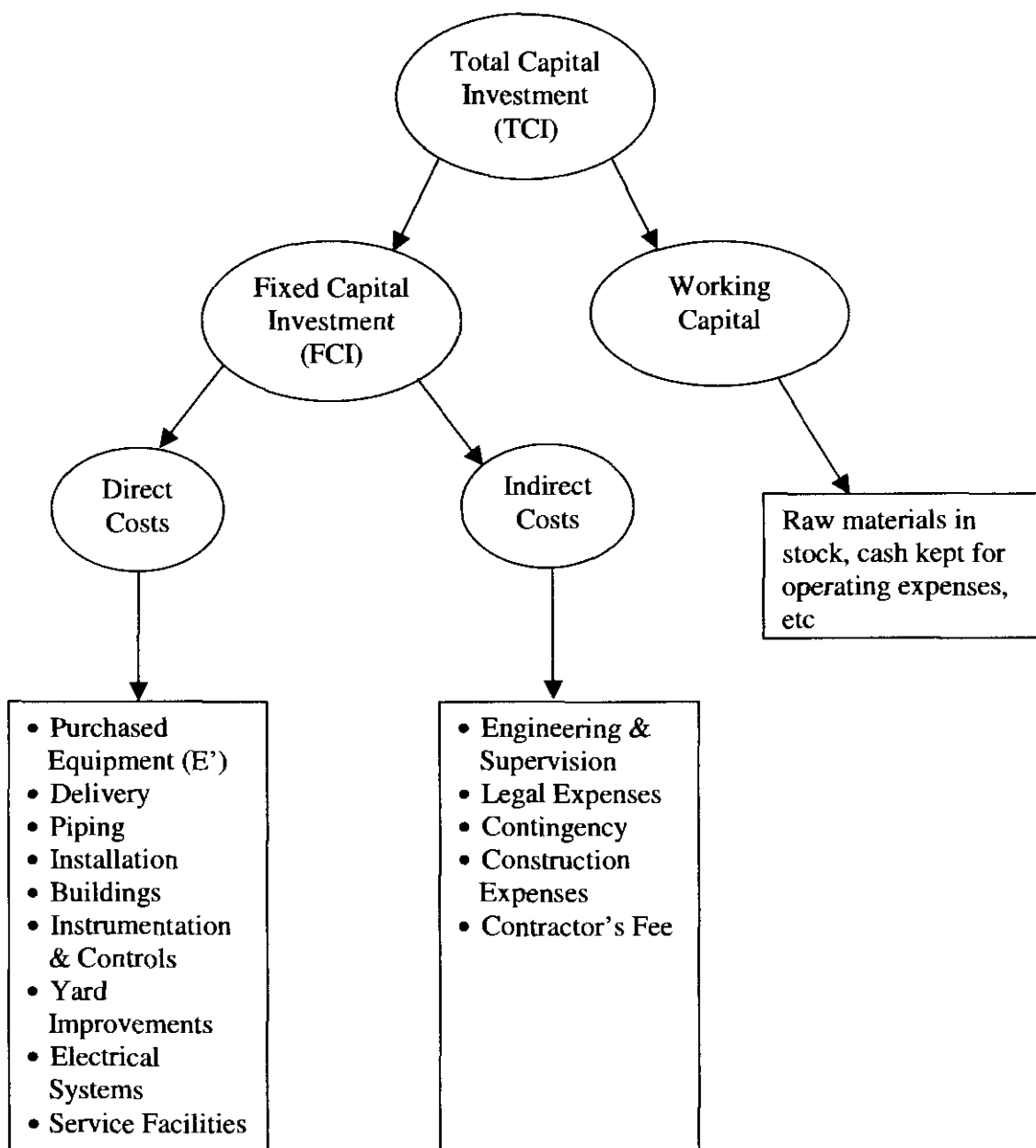


Figure 6. Capital investment cost breakdown.

Purchased equipment costs

By using the mass flow rate outputs from the biodiesel process model (Tapasvi et al., 2005) and the equipment sizing equations available in the literature (Perry et al., 1997; Peters et al., 2003; Saravacos et al., 2002; Sheehan et al., 1998), preliminary equipment

size and specification calculations were performed for determining the purchased equipment costs. Various types of biodiesel production equipment (refer to Figure 2 in Paper 1) are divided into the following nine categories.

1. Heat exchangers – Heaters (A), (G), (N), (O), (R), (U), the reboilers for the distillation columns (V) and (W) and the condenser (X).
2. Mixing tanks with agitator – Mixing tank (B) and refining tank (E)
3. Centrifuges – Centrifuge equipment (C) and (F).
4. Storage tanks/settlers – Horizontal storage vessels (D), (I), (K), (M), (Q), (T), (Y) and (AA) and 10 storage tanks.
5. Jacketed vacuum kettles – Vacuum oil dryer (H) and vacuum ester dryer (S).
6. Jacketed stirred tank reactors – Continuous stirred tank reactors (J), (L) and (Z).
7. Vertical liquid-liquid extraction column – Ester wash column (P).
8. Sieve tray distillation columns – Glycerol-alcohol distillation column (V) and methanol recovery distillation column design (W).
9. Centrifugal pumps – Inline vertical motor centrifugal pumps were assumed between equipment: (B) to (C), (C) to (E), NaOH solution to (E), (Soft water) to (E), (E) to (F), (H) to (I), (I) to (J), methanol to (J) and (L), (K) to (L) and one spare pump.

All the equipment is the same as selected by Tapasvi et al., 2005, except the super-heated glycerol-alcohol stripping column (V). Due to the lack of sizing equations in the literature for a super-heated steam stripping column and the high cost similarity between a stripping column and a distillation column, a regular glycerol-alcohol distillation column was chosen for the purpose of sizing and cost estimation in this study.

The equipment sizing calculations are performed separately for two canola biodiesel production plant capacities—5 MGY and 30 MGY—to analyze the economies of scale for biodiesel production plants. The detailed equipment sizing calculations together with the assumptions made for individual equipment are available in the Appendix A.

Table 6 summarizes all the calculated equipment specifications required to determine the total purchased equipment cost (E'). Table 7 summarizes the estimated equipment costs for both biodiesel production plants.

Based on these equipment specifications, the purchased equipment costs are determined from the cost values available from the similar past purchase order database (Peters et al., 2003). The latest available Chemical Engineering (CE) plant cost index, 457.4 for August 2004 (California consumer products regulation, 2004), was utilized to correct these cost values for purchasing new equipment, as these values were obtained in January 2002 (CE index = 390.4) using the following equation: Present cost = original cost * (457.4/390.4) (1).

Also, the six-tenths factor rule was used for the cost estimate of the equipment (H), (J), (L), (P) and storage tanks for the 30 MGY biodiesel production plant, as these costs were unavailable in the past purchase order database (Peters et al., 2003).

Cost of equipment 'a' = (cost of equipment 'b') * $X^{0.6}$ (2) where the capacity of equipment 'a' is X times that of equipment 'b'.

Other direct costs

Other direct costs, such as equipment delivery costs, equipment installation costs, instrumentation and controls costs, piping costs, and electrical systems costs, were

estimated as a fixed fraction of the total purchased equipment cost (E') and are summarized in Table 8 (Peters et al., 2003).

Table 6. Summary of calculated equipment specifications*

Equipment ID/Code	Equipment Type	Material of Construction	Pressure	Specifications 5 MGY	Specifications 30 MGY
A	Double pipe HE	Carbon steel	Up to 4135 kPa	SA = 0.62 m ²	SA = 3.73 m ²
B	Mixing tank with agitator	Carbon steel	Up to 101 kPa	V = 3 m ³	V = 17.42 m ³
C	Tubular bowl centrifuge	Carbon steel	Up to 101 kPa	PC = 10.2 KW	PC = 60.9 KW
D	Horizontal storage vessel	Carbon steel	Up to 101 kPa	D = 0.5 m, L = 1.5 m	D = 0.5 m, L = 4.48 m
E	Mixing tank with agitator	Carbon steel	Up to 101 kPa	V = 3.30 m ³	V = 19.60 m ³
F	Tubular bowl centrifuge	Carbon steel	Up to 101 kPa	PC = 10.2 KW	PC = 60.9 KW
G	Double pipe HE	Carbon steel	Up to 4135 kPa	SA = 0.32 m ²	SA = 1.93 m ²
H	Jacketed kettle	Cast-iron	Up to 101 kPa	V = 2.8 m ³	V = 16.70 m ³
I	Horizontal storage vessel	Carbon steel	Up to 101 kPa	D = 1.0 m, L = 6.8 m	D = 2.0 m, L = 10.2 m
J	Jacketed and stirred reactor	Stainless steel	Up to 345 kPa	V = 7 m ³	V = 41.8 m ³
K	Horizontal storage vessel	Carbon steel	Up to 101 kPa	D = 2.0 m, L = 8.53 m	D = 3.0 m, L = 22.7 m
L	Jacketed and stirred reactor	Stainless steel	Up to 345 kPa	V = 6.3 m ³	V = 37.6 m ³
M	Horizontal storage vessel	Carbon steel	Up to 101 kPa	D = 2.0 m, L = 7.50 m	D = 3.0 m, L = 20.1 m
N	Double pipe HE	Carbon steel	Up to 4135 kPa	SA = 0.23 m ²	SA = 1.26 m ²
O	Double pipe HE	Carbon steel	Up to 4135 kPa	SA = 0.23 m ²	SA = 0.55 m ²
P	Vertical column	Carbon steel	Up to 101 kPa	Capacity = 2540 Kg/hr, D=2 m, H= 10 m	Capacity = 15200 Kg/hr
Q	Horizontal storage vessel	Carbon steel	Up to 101 kPa	D = 1.0 m, L = 3.60 m	D = 2.0 m, L = 5.3 m
R	Double pipe HE	Carbon steel	Up to 4135 kPa	SA = 0.55 m ²	SA = 3.24 m ²
S	Jacketed kettle	Cast iron	Up to 101 kPa	V = 2.90 m ³	V = 17.1 m ³
T	Horizontal storage vessel	Carbon steel	Up to 101 kPa	D = 1.0 m, L = 2.8 m	D = 2.0 m, L = 4.3 m

Table 6. (Continued)

Equipment Name	Equipment Type	Material of Construction	Pressure	Specifications 5 MGY	Specifications 30 MGY
U	Double pipe HE	Carbon steel	Up to 4135 kPa	SA = 0.23 m ²	SA = 1.28 m ²
V	Sieve tray column	Carbon steel	Up to 101 kPa	D = 0.5 m, H = 3.7 m, 6 stages	D = 1.0 m, H = 3.7 m, 6 stages
W	Sieve tray column	Carbon steel	Up to 101 kPa	D = 1.0 m, H = 12.2 m, 20 stages	D = 2.0 m, H = 12.2 m, 20 stages
X	Double pipe HE	Carbon steel	Up to 4135 kPa	SA = 11.80 m ²	SA = 70.1 m ²
Y	Horizontal storage vessel	Carbon steel	Up to 101 kPa	D = 0.5 m, L = 2.48 m	D = 1.0 m, L = 3.72 m
Z	Jacketed and stirred reactor	Carbon steel	Up to 345 kPa	V = 0.41 m ³	V = 2.5 m ³
AA	Horizontal storage vessel	Carbon steel	Up to 101 kPa	D = 0.5 m, L = 4.1 m	D = 1.0 m, L = 6.2 m
Reboiler for V	Double pipe HE	Carbon steel	Up to 4135 kPa	SA = 9.3 m ²	SA = 55.2 m ²
Reboiler for W	Double pipe HE	Carbon steel	Up to 4135 kPa	SA = 2.1 m ²	SA = 12.2 m ²
10 storage tanks	Small, field-erected	Carbon steel	Up to 101 kPa	V = 285 m ³ (each)	V = 1710 m ³ (each)
CP for stream 5	Inline, vertical motor	Cast steel	Up to 1035 kPa	VFR (m ³ /s)* DP (kPa) = 3.0	VFR (m ³ /s)* DP (kPa) = 4.6
CP for stream 9	Inline, vertical motor	Cast steel	Up to 1035 kPa	VFR (m ³ /s)* DP (kPa) = 3.0	VFR (m ³ /s)* DP (kPa) = 4.5
CP for stream 10	Inline, vertical motor	Cast steel	Up to 1035 kPa	VFR (m ³ /s)* DP (kPa) = 3.0	VFR (m ³ /s)* DP (kPa) = 3.0
CP for stream 11	Inline, vertical motor	Cast steel	Up to 1035 kPa	VFR (m ³ /s)* DP (kPa) = 3.0	VFR (m ³ /s)* DP (kPa) = 3.0
CP for stream 12	Inline, vertical motor	Cast steel	Up to 1035 kPa	VFR (m ³ /s)* DP (kPa) = 3.0	VFR (m ³ /s)* DP (kPa) = 5.2
CP for stream 17	Inline, vertical motor	Cast steel	Up to 1035 kPa	VFR (m ³ /s)* DP (kPa) = 3.0	VFR (m ³ /s)* DP (kPa) = 4.4
CP for stream 18	Inline, vertical motor	Cast steel	Up to 1035 kPa	VFR (m ³ /s)* DP (kPa) = 3.0	VFR (m ³ /s)* DP (kPa) = 4.4
CP for streams 20 & 25	Inline, vertical motor	Cast steel	Up to 1035 kPa	VFR (m ³ /s)* DP (kPa) = 3.0	VFR (m ³ /s)* DP (kPa) = 3.0
CP for stream 23	Inline, vertical motor	Cast steel	Up to 1035 kPa	VFR (m ³ /s)* DP (kPa) = 3.0	VFR (m ³ /s)* DP (kPa) = 4.8
Spare CP	Inline, vertical motor	Cast steel	Up to 1035 kPa	VFR (m ³ /s)* DP (kPa) = 3.0	VFR (m ³ /s)* DP (kPa) = 5.2

* A, B, C, D, etc. are the biodiesel production equipment identified in Figure 2 of Paper 1. SA, V, PC, VFR, DP, CP, HE, D, L and H stand for surface area, volume, power consumption, volumetric flow rate, differential pressure, centrifugal pump, heat exchanger, diameter, length, and height, respectively. Refer to Appendix A for detailed sizing calculations.

Table 7. Summary of the estimated purchased equipment cost (August 2004)

Equipment Name	5 MGY (\$ US)	30 MGY (\$ US)
A	1,181	1,324
B	13,736	34,607
C	41,510	122,538
D	1,597	2,895
E	14,441	36,818
F	41,510	122,538
G	1,133	1,270
H	31,644	92,259
I	7,687	20,460
J	69,598	203,932
K	17,836	63,946
L	65,824	192,263
M	17,836	63,946
N	1,110	1,236
O	1,110	1,173
P	45,531	133,147
Q	4,780	12,707
R	1,173	1,312
S	32,111	93,113
T	4,046	11,105
U	1,110	1,237
V - (Vertical column and sieve trays)	13,646	17,674
W- (Vertical column and sieve trays)	43,652	73,455
X	1,425	4,162
Y	2,029	4,890
Z	8,746	23,156
AA	2,732	7,146
Reboiler for V	1,404	4,087
Reboiler for W	1,277	1,427
Centrifugal pump for stream 5	7,638	8,210
Centrifugal pump for stream 9	7,638	8,174
Centrifugal pump for stream 10	7,638	7,638
Centrifugal pump for stream 11	7,638	7,638
Centrifugal pump for stream 12	7,638	8,416
Centrifugal pump for stream 17	7,638	8,140
Centrifugal pump for stream 18	7,638	8,140
Centrifugal pump for streams 20 & 25	7,638	7,638
Centrifugal pump for stream 23	7,638	8,279
Spare centrifugal pump	7,638	8,416
10 storage tanks	496,087	1,453,616
Total (E')	1,063,878	2,884,127

Table 8. Direct costs involved in the biodiesel production plants, million \$/year

	Fraction of E'	5 MGY	30 MGY
Purchased equipment, E'	1.00	1.070	2.890
Equipment delivery'	0.10	0.107	0.289
Purchased equipment installation	0.47	0.553	1.494
Instrumentation & Controls (installed)	0.36	0.424	1.144
Piping (installed)	0.68	0.800	2.162
Electrical systems (installed)	0.11	0.129	0.350
Buildings (including services)	0.18	0.212	0.572
Yard improvements	0.10	0.118	0.318
Service facilities (installed)	0.70	0.824	2.225
Total direct costs	3.70	4.237	11.444

Indirect costs calculations

Indirect costs, such as costs involved in engineering and supervision, construction expenses, legal expenses, contractor's fee, and contingency, are also estimated based on a fixed fraction of the total purchased equipment cost (E') and are summarized in Table 9 (Peters et al., 2003).

Fixed capital investment calculations

Fixed capital investment (FCI) is the sum of direct costs and indirect costs (Peters et al., 2003).

Fixed capital investment for the 5 MGY biodiesel production plant
 = \$ 4.237 million + \$1.695 million = **\$ 5.932 million**

Fixed capital investment for the 30 MGY biodiesel production plant
 = \$11.444 million + \$ 4.578 million = **\$ 16.030 million**

Working capital calculations

The working capital (WC) estimates are based on a fixed fraction (0.89) of the delivered equipment cost (E' + equipment delivery cost) (Peters et al., 2003).

Working capital estimate for the 5 MGY biodiesel production plant
= \$ 1.177 million * 0.89 = **\$ 1.048 million**

Working capital estimate for the 30 MGY biodiesel production plant
= \$ 3.179 million * 0.89 = **\$ 2.830 million**

Total capital investment (TCI) cost calculations

The total capital investment (TCI) is the sum of fixed capital investment (FCI) and the working capital (WC).

Total capital investment cost for the 5 MGY biodiesel production plant
= \$ 5.932 million + \$ 1.048 million = **\$ 6.980 million**

Total capital investment cost for the 30 MGY biodiesel production plant
= \$ 16.030 million + \$ 2.830 million = **\$ 18.900 million**

Table 9. Indirect cost estimation for biodiesel production plants, million \$/year

	Fraction of E'	5 MGY	30 MGY
Engineering and supervision	0.33	0.388	1.049
Construction expenses	0.41	0.483	1.303
Legal expenses	0.04	0.047	0.127
Contractor's fee	0.22	0.259	0.699
Contingency	0.44	0.518	1.399
Total indirect costs	1.44	1.695	4.578

Estimation of revenue

Revenue mainly comes from the sale of the product biodiesel and the co-product glycerol. The total revenue from product sales is the sum of the unit price of each product multiplied by its rate of sales as shown in Table 10 (Peters et al., 2003).

Table 10. Product and co-product cost for both plants

Name of Material	Price, \$/kg	Annual amount for the 5 MGY plant, million kg/y	Annual amount for the 30 MGY plant, million kg/y	Annual value of product from the 5 MGY plant, million \$/y	Annual value of product from the 30 MGY plant, million \$/y
Biodiesel ^{1,2}	0.78	16.584	99.230	12.94	77.40
Glycerol ³	0.75	2.125	12.711	1.60	9.54
Soapstock ³	0.01	3.174	18.991	0.03	0.19
Total annual value of products =				14.57	87.13

¹ Fuel Price Report (2006).

² Deep Green Crystals archives (2006).

³ Fortenbery (2005).

Estimation of total product cost

Total product cost includes the total of all costs of operating the plant, selling the products, recovering the capital investment, and contributing to corporate functions, such as management, research, and development. It is divided into two categories: manufacturing costs and general expenses.

Manufacturing costs

All expenses directly connected with the operation or the physical equipment of a plant are included in the manufacturing costs and are divided into three classifications:

1) variable production costs, 2) fixed charges, and 3) plant overhead costs.

Variable production costs

These costs involve expenditures for raw materials (including transportation, unloading, etc.), direct operating labor, and supervisory and clerical labor directly applied to the manufacturing operation, utilities, plant maintenance and repairs, operating supplies, etc.

1. Raw material costs

Raw material costs for both plants are presented in Table 11.

2. Operating labor, supervision, clerical, and administrative assistance

Estimates made in VanWechel et al., 2002 have been used as a reference to calculate costs in Table 12 for a 5 MGY biodiesel production plant in North Dakota.

The relationship between labor requirements and production rate is not a linear one; a 0.25 power of the capacity ratio when plant capacities are scaled up was used in this study (Peters et al., 2003). Therefore, the labor expenses for a 30 MGY biodiesel plant is as follows:

$$\text{Labor expenses (30 MGY)} = \$ 343,620 * (6)^{0.25} = \$ 537,800$$

3. Utilities costs

Utility costs include the cost for utilities such as process water, electricity, steam, wastewater treatment and solid waste disposal. These costs vary widely depending on the amount needed, plant location and source. Table 13 summarizes the calculated utilities costs for the 5 MGY and 30 MGY biodiesel production plants, respectively.

4. Maintenance and repairs

In the process industries, the total plant cost per year for maintenance and repairs ranges from 2 to 10% of the fixed-capital investment (FCI), and 7% was chosen for this study (Peters et al., 2003). Tables 14 and 15 summarize these costs.

5. Operating supplies

Consumable items such as charts, lubricants, test chemicals, custodial, and similar supplies cannot be considered as raw materials or maintenance and repair materials, and these are classified as operating supplies. The annual cost for these types of supplies is about 15% of the total cost for maintenance and repairs (Peters et al., 2003). Tables 14 and 15 summarize these costs.

Table 11. Raw material costs for both plants. Chemical Market Reporter (December 2005).

Name of Material	Price, \$/kg	Annual amount for the 5 MGY plant, million kg/y	Annual amount for the 30 MGY plant, million kg/y	Annual raw materials cost for the 5 MGY plant, million \$/y	Annual raw materials cost for the 30 MGY plant, million \$/y
Crude canola oil	0.63	17.400	104.120	10.93	65.39
Methanol	0.32	3.590	21.460	1.15	6.87
Sodium methoxide, 25%	1.12	0.765	4.576	0.86	5.13
0.85% phosphoric acid	0.83	0.017	0.089	0.01	0.07
50% NaOH solution	0.32	0.053	0.315	0.02	0.10
35% HCL solution	0.08	0.460	2.730	0.04	0.22
Total annual cost of raw materials =				13.01	77.78

Table 12. Labor cost estimation for the 5 MGY biodiesel production plant. North Dakota Career Resource Network (2004).

Position	Number of Employees	Salary (US \$/yr)
Staff-operators	4	4 * 48,060 = 192,240
Administrator	1	51,802
Lab technician	1	34,612
Support staff	1	24,966
Sales person	1	40,000
Total	8	343,620

Table 13. Utility costs for both plants

Utility	Cost/unit	Annual requirement for 5 MGY plant	Annual requirement for 30 MGY plant	Annual utility cost for 5 MGY plant, million \$/yr	Annual utility cost for 30 MGY plant, million \$/year
Process water ¹	\$1.06/1000 kg	7.2*10 ⁶ kg/yr	4.3*10 ⁷ kg/yr	0.0077	0.0458
Electricity ²	\$0.06/kWh	4.15*10 ⁵ kWh	2.49*10 ⁶ kWh	0.0249	0.1494
Saturated steam, 150 psi, 180°C ²	\$0.0071/kg	1.78*10 ⁷ kg/yr	1.06*10 ⁸ kg/yr	0.1260	0.7536
Waste water ²	\$0.15/kg	4.02*10 ⁶ kg/yr	2.41*10 ⁷ kg/yr	0.6040	3.6102
Waste disposal ³	\$0.035/kg	3.18*10 ⁵ kg/yr	1.90*10 ⁶ kg/yr	0.0112	0.0664
TOTAL				0.7738	4.6254

¹ VanWechel et al. (2002).

² Peters et al. (2003).

³ Fortenbery (2005).

Fixed charges

Costs that change little or not at all with the amount of production are designated as fixed costs or fixed charges. These include primarily the following costs.

1. Depreciation costs

The depreciation estimates came from the IRS publication 946 using the straight-line general depreciation method. The building is considered as a

nonresidential real property with a 25-year recovery period, and the biodiesel production equipment, including the storage tanks, is considered as a vegetable oil products manufacturing facility with a 10-year recovery period. Therefore, the building cost was multiplied by 4%, while the production equipment cost was multiplied by 10% to get the total depreciation numbers for both 5 million gallon and 30 million gallon biodiesel production costs (US Department of Treasury, IRS Publication, 2005).

Depreciation cost estimate for the 5 MGY biodiesel production plant
 $= (\$ 3.083 \text{ million} * 0.10) + (\$ 0.212 \text{ million} * 0.04) = \$ 316,780$

Depreciation cost estimate for the 30 MGY biodiesel production plant
 $= (\$ 8.329 \text{ million} * 0.10) + (\$ 0.572 \text{ million} * 0.04) = \$ 855,780$

Table 14. Total product cost for a 5 MGY biodiesel production plant

Item	Factor	Basis	Basis cost, million \$/y	Cost, million \$/y	Cost, \$/gallon of product
Raw materials				13.010	2.602
Operating labor and supervision				0.344	0.069
Utilities				0.774	0.155
Maintenance and repairs	0.07	of FCI	5.930	0.420	0.084
Operating supplies	0.15	of maintenance & repair	0.420	0.063	0.013
Insurance	0.01	of FCI	5.930	0.059	0.012
Depreciation				0.317	0.064
Interest				0.140	0.028
Plant overhead, general	0.6	of labor, supervision and maintenance	0.764	0.458	0.092
Distribution & marketing	0.015	of TCI	6.980	0.110	0.022
TOTAL PRODUCT COST/YEAR				15.97	3.195

2. Interest costs

Interest costs were calculated using Farm Credit Services rates. A 10-year fixed interest rate of 8% with 40% equity was used to calculate the costs for the 5 and 30 MGY biodiesel production plants (VanWechel et al., 2002). Tables 14 and 15 will summarize these costs.

3. Property insurance

Normally, the property insurance charges are about 1% of the fixed-capital investment per year (Peters et al., 2003). Tables 14 and 15 summarize these costs.

Table 15. Total product cost for a 30 MGY biodiesel production plant

Item	Factor	Basis	Basis cost, million \$/y	Cost, million \$/y	Cost, \$/ gallon of product
Raw materials				77.780	2.593
Operating labor and supervision				0.538	0.018
Utilities				4.625	0.154
Maintenance and repairs	0.07	of FCI	6.030	1.120	0.037
Operating supplies	0.15	Of maintenance & repair	1.120	0.168	0.006
Insurance	0.01	of FCI	16.030	0.160	0.005
Depreciation				0.856	0.029
Interest				0.378	0.013
Plant overhead, general	0.6	of labor, supervision and maintenance	1.658	0.995	0.033
Distribution & marketing	0.015	of TCI	18.900	0.290	0.010
TOTAL PRODUCT COST/YEAR				86.87	2.898

Plant overhead costs

The expenditures required for routine plant services are included in plant overhead costs. Non-manufacturing machinery, equipment, and buildings are necessary for many of the general plant services, and the fixed charges and direct costs for those items are part of the plant overhead costs. The plant overhead cost for process plants is about 60% of the total expenses for operating labor, supervision, and maintenance (Peters et al., 2003). Tables 14 and 15 summarize these costs.

General expenses

In addition to the manufacturing costs, other general expenses such as distribution and marketing costs are involved in the operations of a company. These expenses are incurred in the process of selling and distributing the various products. For a biodiesel production plant, these costs are around 1.5% of the total capital investment (TCI) (VanWechel et al., 2002). Tables 14 and 15 summarize these costs.

Gross profit calculations

The product sales revenue minus total product cost gives the gross profit, also called gross earnings (Peters et al., 2003).

Gross annual profit for a 5 MGY plant (without tax incentive)
 $= \$14.57 - \$15.97 = \text{\$-1.40 million}$

Gross annual profit for a 30 MGY plant (without tax incentive)
 $= \$87.13 - \$86.87 = \text{\$ 0.26 million}$

DISCUSSION

Utilization of process engineering and economic principles

The primary purpose for this economic feasibility study is to provide users with a transparent approach utilizing the process engineering and economics principles for calculating the detailed biodiesel equipment specifications, and performing the capital investment cost analysis and the total biodiesel product cost analysis based on the developed biodiesel process model (Tapasvi et al., 2005). The whole cost analysis is presented in a very user friendly way, and, depending upon the user requirements, the cost data used for this study can easily be changed to get the revised feasibility results.

Capital investment cost analysis

Total direct costs are \$4.237 million for the 5 MGY production plant and \$11.444 million for the 30 MGY production plant. These costs are calculated based on the purchased equipment costs, and cost benefits from the economies of scale for the biodiesel production is very much apparent—as although the capacity ratio for these plants is 1:6, the direct costs ratio is just 1:2.7. Similarly, the indirect cost ratio, fixed capital investment (FCI) ratio, working capital (WC) ratio and the total capital investment (TCI) ratio for these plants is close to 1:2.7, as these are calculated based on a fixed fraction of the direct costs.

Annual product cost analysis

Tables 14 and 15 summarize the total biodiesel product cost per year for both 5 MGY and 30 MGY biodiesel production plants, respectively. This represents the cost to produce biodiesel and is not equal to the biodiesel cost at a retail station. As concluded by the previous biodiesel economic feasibility studies, the raw material costs (primarily the

crude canola oil) represents around 83% of the total product cost for the 5 MGY production plant and around 89% of the total product cost for the 30 MGY production plant. The higher percentage for the 30 MGY is also possible because the study has assumed the same per unit price of the raw materials with no consideration to the quantity purchased, due to the unavailability of this type of cost data. If we analyze the items (other than the raw materials costs) such as operating labor cost, utilities cost, maintenance and repair cost, operating supplies cost, insurance cost, depreciation cost, etc, the cost ratios of these items for both production plants is 1:3.43. These results show considerable economies of scale with respect to these items.

Gross profit analysis

Although the raw materials constitute more than 80% of the total product cost for both capacity biodiesel production plants, the yearly gross profit for the 5 MGY is \$ -1.40 million and for the 30 MGY is \$ + 0.26 million. This indicates considerable economies of scale present in biodiesel production, and investing in the higher capacities biodiesel production plants is worthwhile. The current average United States selling price of \$0.78/kg (\$2.57/gallon) for the biodiesel product, \$0.75/kg for the co-product glycerol and \$0.01/kg for the by-product soapstock has been assumed in the study (Chemical Market Reporter, 2005; Fuel price report, 2006; Deep green crystals archives, 2006; Fortenbery, 2005; VanWechel et al., 2002).

Analyzing the results with the federal biodiesel tax incentive

The biodiesel tax incentive is a federal excise tax credit that brings lower-cost biodiesel to biodiesel consumers. The credit equates to one cent per percentage of biodiesel in a fuel blend made from agricultural products, such as vegetable oils, and one-half cent

per percentage for recycled oils. The incentive is taken at the blender level, meaning petroleum distributors, and is passed on to the consumer.

Based on this biodiesel tax incentive, the B100 (100% pure biodiesel) price will be \$1.00 less than its actual price. This will result in the final biodiesel product cost per gallon to be equal to \$2.195 and \$1.898 from the 5 MGY and 30 MGY production plants, respectively. This will result in a significant change in the gross profit analyses for both plants.

Gross annual profit for a 5 MGY plant (with tax incentive)
= \$14.57 – \$10.98 = **\$3.59 million**

Gross annual profit for a 30 MGY plant (with tax incentive)
= \$87.13 – \$56.94 = **\$30.19 million**

This is an encouraging result as compared to the gross profit without the tax incentive. Also, this tax incentive will reduce the price of B20 (20% biodiesel blend in regular diesel) to be approximately the same price as diesel after incorporating the 20-cent tax incentive, thus greatly reducing this cost barrier. Variables that determine the cost differential do exist, including the price of diesel, but the tax incentive will help close the gap (National Biodiesel Board Tax Incentive Fact Sheet, 2004).

Sensitivity analysis of feedstock prices on the final gross profits

Tables 16 and 17 summarize the effect of changing crude canola oil prices on the final per gallon production costs of various biodiesel blends, i.e., B2, B5, B20, and B100. The basis for the analysis includes a fixed petro-diesel price (“Fuel Price Report,” 2006) and B100 production cost calculations from the present study. The costs shown in these

tables include the tax incentive but do not represent the biodiesel selling prices at a retail station.

Table 16. Sensitivity analysis for the 5 MGY biodiesel production plant (with tax incentive)

Crude canola oil (\$/lb)	B2 (\$/gallon)	B5 (\$/gallon)	B20 (\$/gallon)	B100 (\$/gallon)	Diesel (\$/gallon) (April 2006)
0.150	2.612	2.552	2.256	0.678	2.650
0.180	2.617	2.568	2.322	1.014	2.650
0.210	2.624	2.585	2.388	1.349	2.650
0.240	2.631	2.602	2.458	1.686	2.650
0.270	2.637	2.619	2.524	2.021	2.650
0.300	2.644	2.635	2.590	2.357	2.650

Table 17. Sensitivity analysis for the 30 MGY biodiesel production plant (with tax incentive)

Crude canola oil (\$/lb)	B2 (\$/gallon)	B5 (\$/gallon)	B20 (\$/gallon)	B100 (\$/gallon)	Diesel (\$/gallon) (April 2006)
0.150	2.608	2.544	2.226	0.523	2.650
0.180	2.614	2.559	2.286	0.828	2.650
0.210	2.620	2.575	2.348	1.132	2.650
0.240	2.626	2.590	2.408	1.437	2.650
0.270	2.632	2.605	2.470	1.742	2.650
0.300	2.638	2.620	2.530	2.046	2.650

Economic impacts on the local community

A new biodiesel production plant with either 5 or 30 MGY capacity, would have the following economic impacts on the local North Dakota community in which it is located:

1. It will cost approximately \$6.98 million and \$18.9 million to build and equip the 5 MGY and 30 MGY plants, respectively. These costs represent expenditures for goods and services, most of which will be made in the local

economy. Only some specialized biodiesel production equipment such as distillation columns, centrifuges, etc, might require out-state suppliers.

2. The most significant value of building a new biodiesel plant comes from the impact of spending for operations. A 5 MGY biodiesel plant will spend more than \$15 million, whereas a 30 MGY biodiesel plant will spend more than \$ 87 million, annually on goods and services ranging from canola feedstock to labor and utilities such as water, electricity, and natural gas. Virtually all of these purchases will be made from local suppliers, and every dollar spent on annual operations will circulate several times throughout the entire local economy.
3. As calculated in the present study, the 5 and 30 MGY biodiesel production plants will support the creation of as many as 8 and 12 permanent new jobs respectively, throughout the entire North Dakota state economy.
4. Setting up a biodiesel production plant in North Dakota will also generate additional revenue for local canola farmers by increasing canola demand. This effect will be similar to the ethanol production case where, in most circumstances, the corn prices were increased by around 5 to 10 cents per bushel as a result of ethanol production. For every 100 acres of corn produced at the national average yield, this translates into as much as \$1,350 of incremental new revenue (Urbanchuk and Kapell, 2002). The effects on canola production as a result of a biodiesel production plant are expected to be similar.

Limitations of the study

Following are the limitations of the present study:

1. This study has not included transportation costs for the biodiesel production plants, as these costs constitute a very small percentage of the total product costs. For example, in the economic feasibility study conducted by VanWechel et al. (2002), the transportation costs are only 2.2% of the total product costs. Including these costs will require a detailed transportation analysis, which is outside the scope of this study.
2. This study has assumed that sufficient market demand is available for the co-product glycerol sales, but a detailed analysis is required to support this assumption.
3. The total capital investment (TCI) calculations are based on the “study estimate,” also referred to as the “factored estimate.” The basis of these types of estimates is the knowledge of major items of equipment involved in the production process and these estimates include a probable accuracy of up to $\pm 30\%$.
4. Fundamental questions exist about the canola availability for biodiesel production: Will there be enough canola available to satisfy feed, food, and export demand, as well as growing demand for biodiesel? How much canola can go to biodiesel without significantly disrupting other markets? These were outside the scope of the present study.
5. The developed biodiesel process model includes various assumptions that were made for performing mass and energy balance calculations. Even though these

assumptions come from authoritative references, some assumptions may not represent the actual industrial conditions for biodiesel production.

CONCLUSION

Considerable economies of scale are present for biodiesel production in the capital investment cost and the total annual biodiesel cost except the raw material costs. As the raw material costs constitute more than 80% of the total product cost, it is very important to look for cheaper raw materials for biodiesel production in United States. This study has assumed the pure biodiesel selling price (B100) of \$2.57/gallon for calculating the annual gross profits, which is 8 cent lower than the regular petro-diesel price of 2.65/gallon for the month of April 2006. As the petro-diesel prices are increasing, biodiesel has become very competitive and the cost barrier that existed in the past has diminished between the two fuels. Also, with the inclusion of biodiesel tax incentives, the price of 2% (B2), 5% (B5) and even 20% (B20) biodiesel blends with regular diesel is almost at par or less than the petro-diesel prices (refer tables 16 and 17). As shown by this study, to further increase the profits from biodiesel production, it will be worthwhile to invest in a 30 MGY production plant rather than a 5 MGY production plant due to the economies of scale factor in biodiesel production.

REFERENCES

- “Biodiesel Bill Becomes Law in Minnesota.” March 29, 2002. *Biodiesel Bulletin*. A Monthly Newsletter of the National Biodiesel Board.
- California consumer products regulation. Chapter-8. 2004. Retrieved September 15, 2005, from <http://www.arb.ca.gov/regact/conprod/ch8.pdf>
- Canola Council of Canada. 2005. Canadian canola production. Retrieved November 8, 2005, from <http://www.canola-council.org/acreageyields.html>
- Chemical Market Reporter. December 2005. Retrieved November 17, 2005, from <http://www.chemicalmarketreporter.com>
- Coltrain, D. Biodiesel: Is it worth considering? Unpublished manuscript presented at Risk and Profit Conference, Kansas State University, Manhattan, KS, p. 19, April 15-16, 2002.
- Conrad, K. United States Senator of North Dakota. March 22, 2005. Retrieved January 14, 2006, from <http://conrad.senate.gov/~conrad/releases/05/03/2005322908.html>
- Deep Green Crystals Archives. April 2006. National average biodiesel price. Retrieved April 6, 2006, from <http://www.martinandalex.com/blog/archives/2006/04/index.html>
- English, B., Jensen, K., and Menard, J. Frazier, Barnes and Associates. 2002. Economic feasibility of producing biodiesel in Tennessee. Agri-industry modeling and analysis group, Department of Agricultural Economics, University of Tennessee, Memphis, Tennessee.
- Fortenberry, T.R. 2005. Biodiesel feasibility study: An evaluation of biodiesel feasibility in Wisconsin. Staff Paper No. 481. Department of Agricultural & Applied Economics. University of Wisconsin-Madison.

Frazier Barnes & Associates. 2003. Mississippi statewide biodiesel feasibility study report. Memphis, TN: Prepared for The Mississippi Biomass Council.

Fuel price report. April 2006. Average national petro-diesel price. Retrieved April 5, 2006, from <http://publicservice.vermont.gov/pub/fuel-price-report/06april.pdf>

Galster, J. 2004. Info-TAX motor fuel tax. North Dakota Office of State Tax Commissioner. Bismarck, ND.

Mag, T. 2005. Canola seed and oil processing. Canola Council of Canada. Retrieved July 14, 2004, from <http://www.canola-council.org/PDF/Oilprocessing.pdf#zoom=100>

Minnesota mandate press release "Minnesota diesel now contains two percent biodiesel statewide." September 29, 2005. NBB News. Retrieved January 7, 2006, from http://www.biodiesel.org/resources/pressreleases/gen/20050929_MN_Mandate_Implemented.pdf

Minnesota Soybean Council. Friday focus archive. Retrieved January 5, 2006, from <http://www.mnsoybean.org/News/FridayFocusArticles/BiodieselLegislation.cfm>

National Biodiesel Board tax incentive fact sheet. 2004. Retrieved January 9, 2006, from http://www.bioenergycolorado.com/PDF-Files/20041022_Tax_Incentive_Fact_Sheet.pdf

National Agricultural Statistics Service. 2005. Retrieved July 15, 2006, from <http://www.nass.usda.gov/nd/canola71.pdf>

North Dakota Career Resource Network. 2004. Employment and wages analysis. Retrieved October 6, 2005, from http://www.state.nd.us/jsnd/docs/lmi/cew/Qtr_4_2004.pdf

North Dakota Office of Governor. October 5, 2005, press release. Retrieved January 15, 2006, from <http://www.governor.state.nd.us/media/news-releases/2005/10/051004a.html>

- North Dakota Office of Governor. October 22, 2005, press release. Retrieved January 12, 2006, from <http://www.governor.state.nd.us/media/news-releases/2005/04/050422.html>
- Perry, R.H., and Green, D.W. 1997. *Perry's Chemical Engineers' Handbook*. New York: McGraw Hill.
- Peters, M.S., Timmerhaus, K.D. & West, R.E. 2003. *Plant Design and Economics for Chemical Engineers*. 5th Edition. New York: McGraw Hill.
- Pure biodiesel price. June 2005. Dieselstop.com forums. Retrieved December 20, 2005, from <http://forums.thedieselstop.com>
- Saravacos, G.D., & Kostaropoulos, A.E. 2002. *Handbook of Food Processing Equipment*. New York: Kluwer Academic/Plenum Publishers.
- "Senate Approves Energy Bill With Biodiesel Provisions." May 1, 2002. *Biodiesel Bulletin*. A Monthly Newsletter of the National Biodiesel Board.
- Sheehan, J., Camobreco, V., Duffield, J., Graboski, M. & Shapouri, H. 1998. Life cycle inventory of biodiesel and petroleum diesel for use in an urban bus. NREL/SR-580-24089. Golden, CO: National Renewable Energy Laboratory.
- Shumaker, G.A., J. Mckissick, C. Ferland, B. Doherty. 2000. A study on the feasibility of biodiesel production in Georgia. University of GA Center for Agribusiness and Economic Development.
- Soygrowers newsroom. Soyradio. January 2006. Retrieved June 4, 2006 from <http://www.soygrowers.com/newsroom/soyradio/duncanson-010306.htm>
- Tapasvi, D., Wiesenborn, D. & Gustafson, C. 2005. Process model for biodiesel production from various feedstocks. ASAE Paper No. FPE 5688. St Joseph, MI: ASAE.

- Urbanchuk, J.M., & Kapell, J. 2002. Ethanol and the local community. AUS Consultants and SJH & Company. Mt. Laurel, New Jersey.
- U.S. Department of Energy, Energy information administration, petroleum supply monthly (June 2003), DOE/EIA-0109 (2003/06), 2-7.
- U.S. Department of Energy. Energy pact act of 2005 (EPACT). Retrieved January 17, 2005, from <http://www.energy.gov/taxbreaks.htm>
- U.S. Department of Treasury, IRS. Publication 946 (2005), How to depreciate property for use in 2005 returns. Retrieved August 6, 2006, from <http://www.irs.gov/pub/irs-pdf/p946.pdf>
- Van Dyne, D.L. 1992. Biodiesel production potential from industrial rapeseed in the southeastern U.S. Tennessee Valley Authority, the South Regional Biomass Energy Program.
- VanWechel, T, Gustafson, C.R. & Leistritz, E.L. 2002. Economic feasibility of biodiesel production in North Dakota. Agribusiness and Applied Economics Report no. 505, Agricultural Experiment Station, NDSU, Fargo, ND.
- Zhang, Y., Dube, M.A., McLean, D.D. & Kates, M. 2003. Biodiesel production from waste cooking oil: 2. Economic assessment and sensitivity analysis. Bioresour. Technol.

APPENDIX A

EQUIPMENT SELECTION AND SIZING

Heat exchangers

Equation used

$$MC_p\Delta T = UA\Delta T_{lmt\delta} \quad (1)$$

where,

M = Mass flow rate, C_p = Specific heat at constant pressure, ΔT = Temperature difference between the incoming and outgoing streams of the fluid being heated, U = Overall heat transfer coefficient, A = Heat exchanger surface area and $\Delta T_{lmt\delta}$ = Log mean temperature difference.

Heater (A)

Assumptions

1. Double pipe heat exchanger with carbon steel tube and shell construction. Pressure – up to 4135 KPa (Peters et al., 2003)
2. Heating medium is saturated steam at 180°C
3. Temperature of stream 39 is 20°C
4. Temperature of stream 40 is 70°C
5. Mass flow rates (M) for stream 39 = stream 40

Sizing calculations

$$M_{5MGY} = 2175 \text{ kg/hr and } M_{30MGY} = 13015 \text{ kg/hr}$$

$$\Delta T_{lmt\delta} = [(180-20) - (180-70)] / [\ln \{ (180-20)/(180-70) \}] = 133.5 \text{ }^\circ\text{C}$$

$$C_p = 1.913 \times 10^3 \text{ J/Kg/}^\circ\text{C}$$

$$\Delta T = (70-20)^{\circ}\text{C} = 50^{\circ}\text{C}$$

$$U = 800 \text{ W/m}^2/\text{C} \text{ (Peters et al., 2003)}$$

Plugging these values in equation (1) and calculating for the surface area A, we get

For the 5 MGY biodiesel production plant

$$\text{Heater surface area } A = 0.54 \text{ m}^2$$

Including a 15% oversize factor, the final surface area required is $A' = 0.62 \text{ m}^2$.

For the 30 MGY biodiesel production plant

$$\text{Heater surface area } A = 3.24 \text{ m}^2$$

Including a 15% oversize factor, the final surface area required is $A' = 3.73 \text{ m}^2$.

Heater (G)

Assumptions

1. Double pipe heat exchanger with carbon steel tube and shell construction. Pressure – up to 4135 KPa (Peters et al., 2003)
2. Heating medium is saturated steam at 180°C
3. Temperature of stream 39 is 20°C
4. Temperature of stream 40 is 70°C
5. Mass flow rates (M) for stream 14 = stream 15

Sizing calculations

$$M_{5\text{MGY}} = 2100.7 \text{ kg/hr and } M_{30\text{MGY}} = 12570.2 \text{ kg/hr}$$

$$\Delta T_{\text{lmtd}} = [(180-70) - (180-90)] / [\ln\{(180-70)/(180-90)\}] = 99.7^{\circ}\text{C}$$

$$C_p = 1.915 \times 10^3 \text{ J/Kg/C}$$

$$\Delta T = (90-70)^{\circ}\text{C} = 20^{\circ}\text{C}$$

$$U = 800 \text{ W/m}^2/\text{C} \text{ (Peters et al., 2003)}$$

Plugging these values into equation (1) and calculating for the surface area A, we get

For the 5 MGY biodiesel production plant

Heater surface area $A = 0.28 \text{ m}^2$

Including a 15% oversize factor, the final surface area required is $A' = 0.32 \text{ m}^2$.

For the 30 MGY biodiesel production plant

Heater surface area $A = 1.68 \text{ m}^2$

Including a 15% oversize factor, the final surface area required is $A' = 1.93 \text{ m}^2$.

Heater (N)

Assumptions

1. Double pipe heat exchanger with carbon steel tube and shell construction. Pressure – up to 4135 KPa (Peters et al., 2003)
2. Heating medium is saturated steam at 180°C
3. Temperature of stream 39 is 20°C
4. Temperature of stream 40 is 70°C
5. Mass flow rates (M) for stream 28 = stream 29

Sizing calculations

$M_{5\text{MGY}} = 2130.0 \text{ kg/hr}$ and $M_{30\text{MGY}} = 12742.0 \text{ kg/hr}$

$\Delta T_{\text{lntd}} = [(180-55) - (180-70)] / [\ln \{ (180-55) / (180-70) \}] = 117.4^\circ\text{C}$

$C_p = 1.930 \times 10^3 \text{ J/Kg/C}$

$\Delta T = (70-55)^\circ\text{C} = 15^\circ\text{C}$

$U = 800 \text{ W/m}^2/\text{C}$ (Peters et al., 2003)

Plugging these values into equation (1) and calculating for the surface area A, we get

For the 5 MGY biodiesel production plant

Heater surface area $A = 0.185 \text{ m}^2$

Including a 15% oversize factor, the final surface area required is $A' = 0.23 \text{ m}^2$.

For the 30 MGY biodiesel production plant

Heater surface area $A = 1.10 \text{ m}^2$

Including a 15% oversize factor, the final surface area required is $A' = 1.26 \text{ m}^2$.

Heater (O)

Assumptions

1. Double pipe heat exchanger with carbon steel tube and shell construction. Pressure – up to 4135 KPa (Peters et al., 2003)
2. Heating medium is saturated steam at 180°C
3. Temperature of stream 39 is 20°C
4. Temperature of stream 40 is 70°C
5. Mass flow rates (M) for stream 30 = stream 31

Sizing calculations

$M_{5\text{MGY}} = 409.03 \text{ kg/hr}$ and $M_{30\text{MGY}} = 2447.6 \text{ kg/hr}$

$\Delta T_{\text{lmtd}} = [(180-20) - (180-70)] / [\ln\{(180-20)/(180-70)\}] = 133.5^\circ\text{C}$

$C_p = 4.186 \times 10^3 \text{ J/Kg/C}$

$\Delta T = (70-20)^\circ\text{C} = 50^\circ\text{C}$

$U = 2250 \text{ W/m}^2/\text{C}$ (Peters et al., 2003)

Plugging these values into equation (1) and calculating for the surface area A, we get

For the 5 MGY biodiesel production plant

Heater surface area $A = 0.08 \text{ m}^2$

Including a 15% oversize factor, the final surface area required is $A' = 0.09 \text{ m}^2$. But the minimum possible (cost effective) surface area for a heat exchanger is $A'' = 0.23 \text{ m}^2$.

Therefore, we will use the final surface area $A'' = 0.23 \text{ m}^2$ for cost calculations.

For the 30 MGY biodiesel production plant

Heater surface area $A = 0.48 \text{ m}^2$

Including a 15% oversize factor, the final surface area required is $A' = 0.55 \text{ m}^2$.

Heater (R)

Assumptions

1. Double pipe heat exchanger with carbon steel tube and shell construction. Pressure – up to 4135 KPa (Peters et al., 2003)
2. Heating medium is saturated steam at 180°C
3. Temperature of stream 39 is 20°C
4. Temperature of stream 40 is 70°C
5. Mass flow rates (M) for stream 35 = stream 36

Sizing calculations

$M_{5\text{MGY}} = 2074.8 \text{ kg/hr}$ and $M_{30\text{MGY}} = 12415.2 \text{ kg/hr}$

$\Delta T_{\text{limd}} = [(180-60) - (180-95)] / [\ln\{(180-60)/(180-95)\}] = 101.5^\circ\text{C}$

$C_p = 1.890 \times 10^3 \text{ J/Kg/C}$

$\Delta T = (95-60)^\circ\text{C} = 35^\circ\text{C}$

$U = 800 \text{ W/m}^2/\text{C}$ (Peters et al., 2003)

Plugging these values into equation (1) and calculating for the surface area A, we get

For the 5 MGY biodiesel production plant

Heater surface area $A = 0.47 \text{ m}^2$

Including a 15% oversize factor, the final surface area required is $A' = 0.55 \text{ m}^2$.

For the 30 MGY biodiesel production plant

Heater surface area $A = 2.81 \text{ m}^2$

Including a 15% oversize factor, the final surface area required is $A' = 3.24 \text{ m}^2$.

Heater (U)

Assumptions

1. Double pipe heat exchanger with carbon steel tube and shell construction. Pressure – up to 4135 KPa (Peters et al., 2003)
2. Heating medium is saturated steam at 180°C
3. Temperature of stream 39 is 20°C
4. Temperature of stream 40 is 70°C
5. Mass flow rates (M) for stream 39 = stream 40

Sizing calculations

$M_{5\text{MGY}} = 940.0 \text{ kg/hr}$ and $M_{30\text{MGY}} = 5620.0 \text{ kg/hr}$

$\Delta T_{\text{lmtd}} = [(180-20) - (180-65)] / [\ln\{(180-20)/(180-65)\}] = 136.3^\circ\text{C}$

$C_p = 3.202 \times 10^3 \text{ J/Kg/C}$

$\Delta T = (65-20)^\circ\text{C} = 45^\circ\text{C}$

$U = 1500 \text{ W/m}^2/\text{C}$ (Saravacos et al., 2002)

Plugging these values into equation (1) and calculating for the surface area A, we get

For the 5 MGY biodiesel production plant

Heater surface area $A = 0.185 \text{ m}^2$

Including a 15% oversize factor, the final surface area required is $A' = 0.213 \text{ m}^2$. But the minimum possible (cost effective) surface area for a heat exchanger is $A'' = 0.23 \text{ m}^2$.

Therefore, we will use the final surface area $A'' = 0.23 \text{ m}^2$ for cost calculations.

For the 30 MGY biodiesel production plant

Heater surface area $A = 1.10 \text{ m}^2$

Including a 15% oversize factor, the final surface area required is $A' = 1.28 \text{ m}^2$.

Note: The sizing calculations for the reboilers of distillation columns (V) and (W) and for the condenser (X) of the distillation column (W) are explained after the sizing calculations for these columns, later in this section.

Mixing tanks with agitator

Equations used:

Filled volume of the tank (V) = Residence time (τ) * Volume flow rate (F) (2)

Mixing Tank (B)

Assumptions (Peters et al., 2003)

1. Mixing tank with agitator and carbon steel as construction material
2. Residence time of 30 minutes
3. Fill factor = 0.5
4. 20% oversize factor

Sizing calculations

$$\rho_{\text{canola oil}} = 906 \text{ kg/m}^3$$

Volume calculations for the 5 MGY biodiesel production plant

$$M_{\text{total}} = M_{\text{stream2}} + M_{\text{stream3}} + M_{\text{stream4}} = 2175.0 + 2.18 + 20.39 \text{ kg/hr} = 2197.6 \text{ kg/hr}$$

$$F_{\text{total}} = 2197.6 / (3600 * 906) = 6.74 * 10^{-4} \text{ m}^3/\text{s}$$

From (2), we get

$$V = 30 \times 60 \times 6.74 \times 10^{-4} = 1.22 \text{ m}^3$$

Actual volume (V'') can be calculated from the fill factor of 0.5,

$$1.22 / V' = 0.5$$

$$V' = 2.50 \text{ m}^3$$

With 20% oversize factor, $V'' = 3.0 \text{ m}^3$

Volume calculations for the 30 MGY biodiesel production plant

$$M_{\text{total}} = M_{\text{stream2}} + M_{\text{stream3}} + M_{\text{stream4}} = 13015.0 + 13.02 + 122.02 \text{ kg/hr} = 13150.0 \text{ kg/hr}$$

$$F_{\text{total}} = 13150.0 / (3600 \times 906) = 4.03 \times 10^{-3} \text{ m}^3/\text{s}$$

From (1), we get

$$V = 30 \times 60 \times 4.03 \times 10^{-3} = 7.26 \text{ m}^3$$

Actual volume (V'') can be calculated from the fill factor of 0.5

$$7.26 / V'' = 0.5$$

$$V' = 14.52 \text{ m}^3$$

With 20% oversize factor, $V'' = 17.42 \text{ m}^3$.

(E) Refining tank

Assumptions (Peters et al., 2003; Tapasvi et al., 2005)

1. Mixing tank with agitator and carbon steel construction
2. Solution of streams 9, 10 and 11 is 87% canola oil and rest is water
3. Residence time of 30 minutes
4. 20% oversize factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{solution } 9+10+11} = 0.87 \cdot 906 + 0.13 \cdot 1000 = 918 \text{ kg/m}^3$$

Volume calculations for the 5 MGY biodiesel production plant

Mass flow rate of solution (stream 9+10+11) (M) = 2498.3 kg/hr

From (2), we get

$$V = (30 \cdot 60 \cdot 2498.3) / (3600 \cdot 918) = 1.37 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 1.37 / 0.5 = 2.74 \text{ m}^3$$

With 20% oversize factor, we get the final volume (V''),

$$V'' = 3.30 \text{ m}^3$$

Volume calculations for the 30 MGY biodiesel production plant

Mass flow rate of solution (stream 9+10+11) (M) = 14949.1 kg/hr

From (1), we get

$$V = (30 \cdot 60 \cdot 14949.1) / (3600 \cdot 918) = 8.15 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 8.15 / 0.5 = 16.3 \text{ m}^3$$

With 20% oversize factor, we get the final volume (V''),

$$V'' = 19.60 \text{ m}^3$$

Centrifuges

Centrifuges (C) and (F)

Assumptions

1. Tubular bowl centrifuge with carbon steel as construction material (Peters et al., 2003)

2. 100 kg/hr of crude vegetable oil has 0.39 KW of power consumption (Sheehan et al, 1998)
3. 20% oversize factor

Sizing calculations

For the 5 MGY biodiesel production plant

Mass flow rate of the crude oil (M) = 2175.00 kg/hr

Therefore, the power consumption (P) = $0.39 \times 2175.0 / 100 = 8.48$ KW

With 20% oversize factor,

$$P' = 10.2 \text{ KW}$$

For the 30 MGY biodiesel production plant

Mass flow rate of the crude oil (M) = 13015.0 kg/hr

Therefore, the power consumption (P) = $0.39 \times 13015.0 / 100 = 50.76$ KW

With 20% oversize factor,

$$P' = 60.90 \text{ KW}$$

Storage tanks/settlers

Equations used

$$\text{Filled Volume of the vessel (V)} = \text{Residence time } (\tau) * \text{Volume flow rate (F)} \quad (3)$$

$$\text{Volume of a horizontal cylinder} = \pi D^2 L / 4 \quad (4)$$

where, D is the diameter and L is the horizontal length of the cylinder.

Settler (D)

Assumptions (Peters et al., 2003; Tapasvi et al., 2005)

1. Horizontal storage vessel with carbon steel construction. Pressure – up to 101 KPa
2. Stream 6 contains 65.8% canola oil and rest is water

3. Residence time of 60 minutes
4. 15% oversize factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{stream 6}} = 0.658 \cdot 906 + 0.342 \cdot 1000 = 938 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

Mass flow rate of stream 6 (M) = 60.23 kg/hr

From (3), we get

$$V = (60 \cdot 60 \cdot 60.23) / (3600 \cdot 938) = 0.065 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 0.065 / 0.5 = 0.13 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 0.15 \text{ m}^3$$

Assuming diameter = 0.5 m, we get from (4),

L = 0.77 m. But the minimum cost effective length is 1.5 m, therefore we choose the final length of the cylinder (L'),

$$L' = 1.5 \text{ m}$$

For the 30 MGY biodiesel production plant

Mass flow rate of stream 6 (M) = 360.38 kg/hr

From (3), we get

$$V = (60 \cdot 60 \cdot 360.38) / (3600 \cdot 938) = 0.384 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 0.384 / 0.5 = 0.768 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 0.88 \text{ m}^3$$

Assuming diameter = 0.5 m, we get from (4),

$$L = 4.48 \text{ m.}$$

Surge tank (I) (Peters et al., 2003; Tapasvi et al., 2005)

Assumptions

1. Horizontal storage vessel with carbon steel construction. Pressure – up to 101 KPa
2. Stream 17 is 100% canola oil.
3. Residence time of 60 minutes
4. 15% oversize factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{stream 17}} = 906 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

Mass flow rate of stream 17 (M) = 2098.9 kg/hr

From (3), we get

$$V = (60 \times 60 \times 2098.9) / (3600 \times 906) = 2.32 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 2.32 / 0.5 = 4.64 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 5.34 \text{ m}^3$$

Assuming diameter = 1.0 m, we get from (4),

$$L = 6.8 \text{ m.}$$

For the 30 MGY biodiesel production plant

Mass flow rate of stream 17 (M) = 12559.5 kg/hr

From (3), we get

$$V = (60 \times 60 \times 12559.5) / (3600 \times 906) = 13.9 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 13.9 / 0.5 = 27.8 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 32.0 \text{ m}^3$$

Assuming diameter = 2.0 m, we get from (4),

$$L = 10.2 \text{ m.}$$

Settler (K)

Assumptions (Peters et al., 2003; Tapasvi et al., 2005)

1. Horizontal storage vessel with carbon steel construction. Pressure – up to 101 KPa
2. Stream 21 contains 70% canola biodiesel and rest is unreacted canola oil
3. Residence time of 240 minutes
4. 15% oversize factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{stream 21}} = 0.70 \times 874 + 0.30 \times 906 = 884 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

Mass flow rate of stream 21 (M) = 2570.0 kg/hr

From (3), we get

$$V = (240 \times 60 \times 2570.0) / (3600 \times 884) = 11.63 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 11.63/0.5 = 23.2 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 26.8 \text{ m}^3$$

Assuming diameter = 2.0 m, we get from (4),

$$L = 8.53 \text{ m.}$$

For the 30 MGY biodiesel production plant

Mass flow rate of stream 21 (M) = 15375.0 kg/hr

From (3), we get

$$V = (240 \times 60 \times 15375.0) / (3600 \times 884) = 69.6 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 69.6/0.5 = 139.2 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 160.1 \text{ m}^3$$

Assuming diameter = 3.0 m, we get from (4),

$$L = 22.7 \text{ m.}$$

Settler (M) (Peters et al., 2003; Tapasvi et al., 2005)

Assumptions

1. Horizontal storage vessel with carbon steel construction. Pressure – up to 101 KPa
2. Stream 26 contains 92% canola biodiesel, 6% liquid methanol and rest is unreacted oil
3. Residence time of 240 minutes
4. 15% oversize factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{stream 26}} = 0.92*874 + 0.06*792 + 0.02*906 = 870 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

Mass flow rate of stream 26 (M) = 2243.0 kg/hr

From (3), we get

$$V = (240*60*2243.0)/(3600*870) = 10.3 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 10.3 / 0.5 = 20.6 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 23.7 \text{ m}^3$$

Assuming diameter = 2.0 m, we get from (4),

$$L = 7.5 \text{ m.}$$

For the 30 MGY biodiesel production plant

Mass flow rate of stream 26 (M) = 13420.0 kg/hr

From (3), we get

$$V = (240*60*13420.0)/(3600*870) = 61.7 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 61.7 / 0.5 = 123.4 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 142.0 \text{ m}^3$$

Assuming diameter = 3.0 m, we get from (4),

$$L = 20.1 \text{ m.}$$

Settler (Q)

Assumptions (Peters et al., 2003; Tapasvi et al., 2005)

1. Horizontal storage vessel with carbon steel construction. Pressure – up to 101 KPa
2. Stream 33 is mostly canola biodiesel, therefore, $\rho_{\text{stream 33}} = 874 \text{ kg/m}^3$
3. Residence time of 30 minutes
4. 15% overdesign factor
5. Fill factor of 0.5

Sizing calculations

For the 5 MGY biodiesel production plant

Mass flow rate of stream 33 (M) = 2113.7 kg/hr

From (3), we get

$$V = (30 \times 60 \times 2113.7) / (3600 \times 874) = 1.21 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 1.21 / 0.5 = 2.42 \text{ m}^3$$

With 15% overdesign factor, we get the final volume (V''),

$$V'' = 2.8 \text{ m}^3$$

Assuming diameter = 1.0 m, we get from (4),

$$L = 3.6 \text{ m.}$$

For the 30 MGY biodiesel production plant

Mass flow rate of stream 33 (M) = 12648.0 kg/hr

From (3), we get

$$V = (30 \times 60 \times 12648.0) / (3600 \times 874) = 7.24 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 7.24/0.5 = 14.48 \text{ m}^3$$

With 15% overdiseign factor, we get the final volume (V''),

$$V'' = 16.7 \text{ m}^3$$

Assuming diameter = 2.0 m, we get from (4),

$$L = 5.3 \text{ m.}$$

Collecting tank (T)

Assumptions (Peters et al., 2003; Tapasvi et al., 2005)

1. Horizontal storage vessel with carbon steel construction. Pressure – up to 101 KPa
2. Solution of stream 22, 27, 32 and 34—stream 22 constitutes 38% of the total mass flow rates, stream 27 is 12%, stream 32 is 45% and remaining 5% of the solution mass flow rate is stream 34. Stream 22 is 51% glycerol and remaining is methanol, stream 27 is 24% glycerol and remaining is methanol, stream 32 is 13.5% methanol and rest is water, and stream 34 is all water.
3. Residence time of 360 minutes
4. 15% overdiseign factor
5. Fill factor of 0.5

Sizing calculations

$$\begin{aligned} \rho_{\text{solution } 22+27+32+34} &= 0.38*0.51*1260 + 0.38*0.49*792 + 0.12*0.24*1260 + 0.12*0.76*792 \\ &\quad + 0.45*0.135*792 + 0.45*0.865*1000 + 0.05*1000 = 988 \text{ kg/m}^3 \end{aligned}$$

For the 5 MGY biodiesel production plant

Mass flow rate of the solution (stream 22+27+32+34) (M) = 940.0 kg/hr

From (3), we get

$$V = (60*60*940.0)/(3600*988) = 0.95 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 0.95/0.5 = 1.90 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 2.20 \text{ m}^3$$

Assuming diameter = 1.0 m, we get from (4),

$$L = 2.8 \text{ m.}$$

For the 30 MGY biodiesel production plant

Mass flow rate of the solution (stream 22+27+32+34) (M) = 5620.0 kg/hr

From (3), we get

$$V = (60 \times 60 \times 5620.0) / (3600 \times 988) = 5.70 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 5.70/0.5 = 11.4 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 13.2 \text{ m}^3$$

Assuming diameter = 2.0 m, we get using (4),

$$L = 4.3 \text{ m.}$$

Glycerol hold tank (Y)

Assumptions (Peters et al., 2003; Tapasvi et al., 2005)

1. Horizontal storage vessel with carbon steel construction. Pressure – up to 101 KPa
2. Stream 46 is mainly glycerol
3. Residence time of 60 minutes
4. 15% oversize factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{stream 46}} = 1260 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

Mass flow rate of stream 46 (M) = 266.0 kg/hr

From (3), we get

$$V = (60 \times 60 \times 266.0) / (3600 \times 1260) = 0.211 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 0.211 / 0.5 = 0.422 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 0.486 \text{ m}^3$$

Assuming diameter = 0.5 m, we get from (4),

$$L = 2.5 \text{ m.}$$

For the 30 MGY biodiesel production plant

Mass flow rate of stream 46 (M) = 1590.0 kg/hr

From (3), we get

$$V = (60 \times 60 \times 1590.0) / (3600 \times 1260) = 1.27 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 1.27 / 0.5 = 2.54 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 2.92 \text{ m}^3$$

Assuming diameter = 1.0 m, we get from (4),

$$L = 3.72 \text{ m.}$$

Settler (AA)

Assumptions (Peters et al., 2003; Tapasvi et al., 2005)

1. Horizontal storage vessel with carbon steel construction. Pressure – up to 101 KPa
2. Stream 49 is 55% glycerol and rest can be assumed as water
3. Residence time of 60 minutes
4. 15% oversize factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{stream 49}} = 0.55 \cdot 1260 + 0.45 \cdot 1000 = 1143 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

Mass flow rate of stream 49 (M) = 387.0 kg/hr

From (3), we get

$$V = (60 \cdot 60 \cdot 387.0) / (3600 \cdot 1143) = 0.34 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 0.34 / 0.5 = 0.68 \text{ m}^3$$

With 15% oversize factor, we get the final volume (V''),

$$V'' = 0.80 \text{ m}^3$$

Assuming diameter = 0.5 m, we get from (4),

$$L = 4.1 \text{ m.}$$

For the 30 MGY biodiesel production plant

Mass flow rate of stream 49 (M) = 2315.0 kg/hr

From (3), we get

$$V = (60 \cdot 60 \cdot 2315.0) / (3600 \cdot 1143) = 2.1 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 2.1/0.5 = 4.2 \text{ m}^3$$

With 15% overdesign factor, we get the final volume (V''),

$$V'' = 4.83 \text{ m}^3$$

Assuming diameter = 1.0 m, we get from (4),

$$L = 6.2 \text{ m.}$$

Storage tanks for crude canola oil and finished biodiesel storage

Assumptions (Peters et al., 2003; VanWechel et al., 2002)

1. Small field-erected tanks, carbon steel
2. 625,000 gallons of storage for crude canola oil and 83,333 gallons of storage for finished biodiesel are required for a 5 MGY biodiesel production plant (Tamara et al). This is equivalent to around 10 small filed erected storage tanks, each with 75,000 gallon or 285 m³ capacity.
3. The costing for the 30 million gallon will be based on the six-tenths factor rule.

Jacketed vacuum kettles

Equations used

$$\text{Filled volume of the kettle (V)} = \text{Residence time } (\tau) * \text{Volume flow rate (F)} \quad (5)$$

(H) Vacuum oil dryer

Assumptions (Peters et al., 2003; Sheehan et al., 1998; Tapasvi et al., 2005)

1. Cast iron jacketed kettle with attached vacuum system
2. Stream 15 is assumed to be mainly canola oil
3. Residence time of 30 minutes
4. 20% overdesign factor

5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{stream 15}} = 906.0 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

Mass flow rate of stream 15 (M) = 2100.7 kg/hr

From (5), we get

$$V = (30 \times 60 \times 2100.7) / (3600 \times 906.0) = 1.16 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 1.16 / 0.5 = 2.32 \text{ m}^3$$

With 20% oversize factor, we get the final volume (V''),

$$V'' = 2.80 \text{ m}^3$$

For the 30 MGY biodiesel production plant

Mass flow rate of stream 15 (M) = 12570.2 kg/hr

From (5), we get

$$V = (30 \times 60 \times 12570.2) / (3600 \times 906.0) = 6.94 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 6.94 / 0.5 = 13.88 \text{ m}^3$$

With 20% oversize factor, we get the final volume (V''),

$$V'' = 16.70 \text{ m}^3$$

(S) Vacuum ester dryer

Assumptions (Peters et al., 2003; Sheehan et al., 1998; Tapasvi et al., 2005)

1. Cast iron jacketed kettle with attached vacuum system
2. Stream 36 is assumed to be mainly canola biodiesel

3. Residence time of 30 minutes
4. 20% overdiseign factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{stream 15}} = 874.0 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

$$\text{Mass flow rate of stream 36 (M)} = 2074.8 \text{ kg/hr}$$

From (5), we get

$$V = (30 \times 60 \times 2074.8) / (3600 \times 874.0) = 1.19 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 1.19 / 0.5 = 2.38 \text{ m}^3$$

With 20% overdiseign factor, we get the final volume (V''),

$$V'' = 2.90 \text{ m}^3$$

For the 30 MGY biodiesel production plant

$$\text{Mass flow rate of stream 36 (M)} = 12415.2 \text{ kg/hr}$$

From (5), we get

$$V = (30 \times 60 \times 12415.2) / (3600 \times 874.0) = 7.12 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 7.12 / 0.5 = 14.24 \text{ m}^3$$

With 20% overdiseign factor, we get the final volume (V''),

$$V'' = 17.1 \text{ m}^3$$

Jacketed stirred tank reactors

Equations used

$$\text{Filled volume of the vessel (V)} = \text{Residence time } (\tau) * \text{Volume flow rate (F)} \quad (6)$$

CSTR (J)

Assumptions (Peters et al., 2003; Tapasvi et al., 2005)

1. Stainless steel jacketed and stirred reactor. Pressure – up to 345 KPa
2. Solution of streams 18, 19 and 20 is assumed to be 18.5% methanol and 81.5% canola oil
3. Residence time of 60 minutes
4. 20% oversize factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{solution } 18+19+20} = 0.185*792 + 0.815*906 = 885 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

Mass flow rate of solution (stream 18+19+20) (M) = 2572.0 kg/hr

From (6), we get

$$V = (60*60*2572.0)/(3600*885) = 2.91 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 2.91/0.5 = 5.82 \text{ m}^3$$

With 20% oversize factor, we get the final volume (V''),

$$V'' = 7.0 \text{ m}^3$$

For the 30 MGY biodiesel production plant

Mass flow rate of solution (stream 18+19+20) (M) = 15390.0 kg/hr

From (6), we get

$$V = (60 \cdot 60 \cdot 15390.0) / (3600 \cdot 885) = 17.40 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 17.40 / 0.5 = 34.80 \text{ m}^3$$

With 20% oversize factor, we get the final volume (V''),

$$V'' = 41.8 \text{ m}^3$$

CSTR (L)

Assumptions (Peters et al., 2003; Tapasvi et al., 2005)

1. Stainless steel jacketed and stirred reactor. Pressure – up to 345 KPa
2. Solution of streams 23, 24 and 25 is assumed to be 78% canola biodiesel, 14% unreacted canola oil and 8% methanol
3. Residence time of 60 minutes
4. 20% oversize factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{solution}} 18+19+20 = 0.08 \cdot 792 + 0.78 \cdot 874 + 0.14 \cdot 906 = 872 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

Mass flow rate of solution (stream 23+24+25) (M) = 2281.0 kg/hr

From (6), we get

$$V = (60 \cdot 60 \cdot 2281.0) / (3600 \cdot 872) = 2.61 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 2.61 / 0.5 = 5.22 \text{ m}^3$$

With 20% oversize factor, we get the final volume (V''),

$$V'' = 6.3 \text{ m}^3$$

For the 30 MGY biodiesel production plant

Mass flow rate of solution (stream 23+24+25) (M) = 13650.0 kg/hr

From (6), we get

$$V = (60 \cdot 60 \cdot 13650.0) / (3600 \cdot 872) = 15.65 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 15.65 / 0.5 = 31.3 \text{ m}^3$$

With 20% oversize factor, we get the final volume (V''),

$$V'' = 37.6 \text{ m}^3$$

CSTR (Z)

Assumptions (Peters et al., 2003; Tapasvi et al., 2005)

1. Stainless steel jacketed and stirred reactor. Pressure – up to 345 KPa
2. Solution of streams 47 and 48 is assumed to be 70% glycerol and 30% water
3. Residence time of 30 minutes
4. 20% oversize factor
5. Fill factor of 0.5

Sizing calculations

$$\rho_{\text{solution } 47+48} = 0.30 \cdot 1000 + 0.70 \cdot 1260 = 1182.0 \text{ kg/m}^3$$

For the 5 MGY biodiesel production plant

Mass flow rate of solution (stream 47 +48) (M) = 399.0 kg/hr

From (6), we get

$$V = (30 \cdot 60 \cdot 399.0) / (3600 \cdot 1182.0) = 0.17 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 0.17/0.5 = 0.34 \text{ m}^3$$

With 20% oversize factor, we get the final volume (V''),

$$V'' = 0.41 \text{ m}^3$$

For the 30 MGY biodiesel production plant

Mass flow rate of solution (stream 47 +48) (M) = 2385.0 kg/hr

From (6), we get

$$V = (30 \times 60 \times 2385.0) / (3600 \times 1182.0) = 1.01 \text{ m}^3$$

With a 0.5 fill factor, actual volume of the vessel (V'),

$$V' = 1.01/0.5 = 2.02 \text{ m}^3$$

With 20% oversize factor, we get the final volume (V''),

$$V'' = 2.50 \text{ m}^3$$

Vertical liquid-liquid extraction column

Wash column (P)

Assumptions (Tapasvi et al., 2005; Zhang et al., 2003)

1. Vertical sieve tray column with 4 stages and carbon steel construction
2. Counter-current washing
3. Height equivalent of a transfer stage (HETS) = 2.5 m

Sizing calculations

For the 5 MGY biodiesel production plant

Mass flow rate of streams 29 + 30 (M) = 2540.0 kg/hr

Diameter = 2.0 m (scaled from the Zhang et al., 2003 calculations)

Height = 10.0 m

For the 30 MGY biodiesel production plant

Mass flow rate of streams 29 + 30 (M) = 15200.0 kg/hr

Diameter = 4.0 m

Height = 10.0 m

Costs will be determined using the six-tenths factor rule based on the equipment for the 5 MGY biodiesel production plant.

Sieve tray distillation columns

(Peters et al., 2003; Sheehan et al., 1998; Tapasvi et al., 2005)

Equations used

$$\text{Vapor flow rate } V = D(R+1) \quad (7)$$

where, R = reflux ratio

$$A_n = V/V_n \quad (8)$$

A_n is the net column area used in the separation process, V is the volumetric flow rate and V_n is the actual vapor velocity.

$$A_c = A_n/0.85 \quad (9)$$

where, A_c is the cross-sectional area of the distillation column

$$d = (4A_c/\pi)^{0.5} \quad (10)$$

where, d is the distillation column diameter.

Distillation column mass balance equations used:

$$F = D + B \quad (11)$$

where, F is the molar flow rate of the feed, D = molar flow rate of the distillate and B is the molar flow rate of the bottoms of a distillation column.

Methanol/component balance:

$$FX_F = DX_D + BX_B \quad (12)$$

where, X_F is the molar concentration of the component in the feed, X_D is the molar concentration of the component in the distillate and X_B is the molar concentration of the component in the bottoms.

From (11) and (12), we get,

$$D/F = (X_F - X_B)/(X_D - X_B) \quad (13)$$

and,

$$B/F = (X_D - X_F)/(X_D - X_B) \quad (14)$$

$$FH_F + q_r = DH_D + BH_B + q_c \quad (15)$$

where H_F is the enthalpy of the feed, H_D is the enthalpy of the distillate, H_B is the enthalpy of the bottoms, q_r is the reboiler duty for a distillation column reboiler, q_c is the condenser duty for a distillation column condenser.

$$q_r = UA\Delta T_{lmtD} \quad (16)$$

$$q_c = UA\Delta T_{lmtC} \quad (17)$$

Glycerol-alcohol distillation column design (V)

Assumptions (Peters et al., 2005; Perry et al., 1997; Sheehan et al., 1998; Tapasvi et al., 2005)

1. Regular distillation column with saturated steam
2. Sieve tray column, carbon steel with 101 KPa pressure
3. Distillate (D) contains 44.2 % water
4. Bottom contains 80% glycerol
5. 6 tray column
6. Top stage temperature = 92.0°C

7. Bottom stage temperature = 123.00°C
8. Surface tension of the distillate at 92°C = 17.8 dynes/cm
9. 80% of flooding
10. Tray spacing = 0.61 m
11. Downcomer occupies 15% of the cross-sectional area of the column
12. 15% overdesign factor
13. No reflux is required
14. $\rho_{\text{liquid water at } 93^{\circ}\text{C}} = 962 \text{ kg/m}^3$, $\rho_{\text{liquid methanol at } 93^{\circ}\text{C}} = 770 \text{ kg/m}^3$
15. Methanol specific heats – $C_{p,l} = 81.6 \text{ Kj/Kgmol-K}$ and $C_{p,v} = 44.06 \text{ Kj/Kgmol-K}$
16. Latent heat of vaporization of methanol at the boiling point, 64.5°C
= 32256 Kj/Kgmol

Sizing calculations

$$(L/V) * (\rho_v / \rho_L)^{0.5} = (1) * (0.81 / 880)^{0.5} = 0.03$$

where

ρ_L = Average liquid density of the distillate

$$= 0.558 * 962 + 0.442 * 792 = 880 \text{ kg/m}^3$$

ρ_v = density of vapor methanol at 92.0°C

$$= (101.3 * 10^3 * 24.2 * 10^{-3}) / (8.314 * 365) = 0.81 \text{ kg/m}^3$$

Using Peters & Timmerhaus Fig 15.5,

$$C_{sb} = 0.12 \text{ m/s}$$

$$V_{n,f} = 0.12 (17.8/20)^{0.2} * \{ (880 - 0.81) / 0.81 \}^{0.5} = 3.86 \text{ m/s}$$

$$V_n = 0.8 * 2.86 = 3.1 \text{ m/s}$$

$$D_{5MGY \text{ (stream 42)}} = 675.0 \text{ kg/hr}$$

$$D_{30\text{MGY (stream 42)}} = 4035.0 \text{ kg/hr}$$

Incorporating the value of D in equation (7), we get

$$V_{5\text{MGY}} = 0.232 \text{ m}^3/\text{s}$$

$$V_{30\text{MGY}} = 1.39 \text{ m}^3/\text{s}$$

Using equations (8), (9), and (10) for calculating the diameter of the distillation column for the 5 MGY biodiesel production plant

$$A_n = 0.232/3.1 = 0.075 \text{ m}^2$$

$$A_c = 0.075/0.85 = 0.088 \text{ m}^2$$

Assuming a 15% oversize factor,

$$A_c' = 0.11 \text{ m}^2$$

$d = (4 \cdot 0.11 / \pi)^{0.5} = 0.38 \text{ m}$. As the minimum required diameter for a similar off-the shelf distillation column is 0.5 m, we will use that for cost analysis instead of 0.38 m diameter.

$$\text{Height of the column (H)} = 0.61\text{m} \cdot 6 \text{ stages} = 3.7 \text{ m}$$

Using equations (8), (9), and (10) for calculating the diameter of the distillation column for the 30 MGY biodiesel production plant

$$A_n = 1.39/3.1 = 0.45 \text{ m}^2$$

$$A_c = 0.45/0.85 = 0.53 \text{ m}^2$$

Assuming a 15% oversize factor,

$$A_c' = 0.61 \text{ m}^2$$

$d = (4 \cdot 0.61 / \pi)^{0.5} = 0.90 \text{ m}$. As the minimum required diameter for a similar off-the shelf distillation column is 1.0 m, we will use that for cost analysis instead of 0.38 m diameter.

$$\text{Height of the column (H)} = 0.61\text{m} \cdot 6 \text{ stages} = 3.7 \text{ m}$$

Reboiler duty calculations

Feed (F) conditions

Feed is liquid with reference temperature 0°C.

$$\text{Water mole \%} = [(43/18)/\{(43/18) + (31.7/32) + (25.3/92)\}]*100 = 65.3\%$$

Similarly, methanol mole % = 27.1% and Glycerol mole% = 7.6%.

$$\text{Temperature} = 64.5^{\circ}\text{C}$$

$$\text{Now, } H_F = H_{F, W} + H_{F, M} + H_{F, G}$$

or,

$$H_F = 0.653*267.9 + 0.271*81.6*64.5 + 0.076*213.5*64.5 = 2650.0 \text{ Kj/Kgmol}$$

[using steam tables]

Distillate (D) conditions

$$\text{Temperature} = 92.0^{\circ}\text{C}$$

$$\text{Methanol mole \%} = [(45/32)/\{(45/32) + (55/18)\}]*100 = 31.3\%$$

$$\begin{aligned} H_D &= [0.313*\{(81.6*64.5) + (32256.0) + (44.06*(92.0-64.5))\}] + [0.687*(2663.3*18)] \\ &= 45060.0 \text{ Kj/Kgmol} \end{aligned}$$

Bottoms (B) conditions

$$\text{Temperature} = 122.0^{\circ}\text{C}$$

$$\text{Glycerol mole \%} = [(80/92)/\{(80/92) + (20/18)\}]*100 = 43.9\%$$

$$H_B = [0.439*(213.44*122.0)] + [0.561*(505*18)] = 16550.0 \text{ Kj/Kgmol}$$

Sizing calculations for the SMGY biodiesel production plant

$$F = 940.0 \text{ Kg/hr} = 34.4 \text{ Kgmol/hr}$$

$$X_F = 0.271$$

$$X_B = 0.0$$

$$X_D = 0.313$$

where, X = Methanol molar concentration.

From (13) and (14), we get,

$$D = 30.0 \text{ Kgmol/hr}$$

$$B = 4.40 \text{ Kgmol/hr}$$

After plugging the values in (15), we get

$$q_r = 1339000.0 \text{ Kj/hr}$$

Reboiler area calculations

$$U = 800 \text{ W/m}^2/\text{K} \text{ (Peters et al., 2003)}$$

$$\Delta T_{\text{imtd}} = (180.0 - 122.0) = 58.0^\circ\text{C}$$

From (16), we get,

$$A = 8.02 \text{ m}^2$$

With 15% oversize factor, we get,

$$A' = 9.3 \text{ m}^2$$

Sizing calculations for the 30MGY biodiesel production plant

$$F = 5620.0 \text{ Kg/hr} = 206.0 \text{ Kgmol/hr}$$

$$X_F = 0.271$$

$$X_B = 0.0$$

$$X_D = 0.313$$

where, X = Methanol molar concentration.

From (13) and (14), we get,

$$D = 179.0 \text{ Kgmol/hr}$$

$$B = 26.0 \text{ Kgmol/hr}$$

Also, from (15), $FH_F + q_r = DH_D + BH_B + q_c$, we get,

$$q_r = 8011590.0 \text{ Kj/hr}$$

Reboiler area calculations

$$U = 800 \text{ W/m}^2/\text{K} \text{ (Peters et al., 2003)}$$

$$\Delta T_{\text{lmtd}} = (180.0 - 122.0) = 58.0^\circ\text{C}$$

From (16), we get,

$$A = 48.0 \text{ m}^2$$

With 15% overdesign factor, we get,

$$A' = 55.2 \text{ m}^2$$

(W) Methanol recovery distillation column design

Assumptions (Peters et al., 2005; Perry et al., 1997; Sheehan et al., 1998; Tapasvi et al., 2005)

1. Sieve tray column, carbon steel with 101 KPa pressure
2. Distillate (D) contains 0.05 % moisture by mass
3. Bottom contains 0.5% methanol by mass
4. 20-tray column
5. Molar reflux ratio (R) = 6
6. Top stage temperature = 65.0°C
7. Bottom stage temperature = 100.01°C
8. Surface tension of methanol at 65°C = 20 dynes/cm
9. 80% of flooding
10. Tray spacing = 0.61 m
11. Downcomer occupies 15% of the cross-sectional area of the column

12. 70% tray efficiency

13. 15% oversize factor

14. Methanol specific heats – $C_{p,l} = 81.6 \text{ KJ/Kgmol-K}$ and $C_{p,v} = 44.06 \text{ KJ/Kgmol-K}$

15. Latent heat of vaporization of methanol at the boiling point, 64.5°C

$$= 32256 \text{ KJ/Kgmol}$$

Sizing calculations

$$(L/V) \cdot (\rho_v/\rho_L)^{0.5} = (6/7) \cdot (1.154/780)^{0.5} = 0.033$$

where,

ρ_v = density of vapor methanol at 100.01°C

$$= (101.3 \cdot 10^3 \cdot 32 \cdot 10^{-3}) / (8.314 \cdot 338) = 1.154 \text{ kg/m}^3$$

ρ_L = density of liquid methanol at 65°C (780 kg/m^3)

Using Peters & Timmerhaus Fig 15.5,

$$C_{sb} = 0.11 \text{ m/s}$$

$$V_{n,f} = 0.11 (20/20)^{0.2} \cdot \{(780-1.154)/1.154\}^{0.5} = 2.86 \text{ m/s}$$

$$V_n = 0.8 \cdot 2.86 = 2.288 \text{ m/s}$$

$$D_{5\text{MGY}} = 295.0 \text{ kg/hr}$$

$$D_{30\text{MGY}} = 1765.0 \text{ kg/hr}$$

Incorporating the value of D in equation (7), we get

$$V_{5\text{MGY}} = 0.502 \text{ m}^3/\text{s}$$

$$V_{30\text{MGY}} = 3.01 \text{ m}^3/\text{s}$$

Using equations (8), (9), and (10) for calculating the diameter of the distillation column for the 5 MGY biodiesel production plant

$$A_n = 0.502/2.288 = 0.219 \text{ m}^2$$

$$A_c = 0.219/0.85 = 0.258 \text{ m}^2$$

Assuming a 15% overdesign factor,

$$A_c' = 0.297 \text{ m}^2$$

$d = (4 \cdot 0.297 / \pi)^{0.5} = 0.62 \text{ m}$. As the minimum required diameter for a similar off-the shelf distillation column is 1.0 m, we will use that for cost analysis instead of 0.38 m diameter.

Height of the column (H) = $0.61 \text{ m} \cdot 20 \text{ stages} = 12.2 \text{ m}$

Using equations (8), (9), and (10) for calculating the diameter of the distillation column for the 30 MGY biodiesel production plant

$$A_n = 3.01 / 2.288 = 1.32 \text{ m}^2$$

$$A_c = 1.32 / 0.85 = 1.55 \text{ m}^2$$

Assuming a 15% overdesign factor,

$$A_c' = 1.79 \text{ m}^2$$

$d = (4 \cdot 1.79 / \pi)^{0.5} = 1.51 \text{ m}$. As the minimum required diameter for a similar off-the shelf distillation column is 2.0 m, we will use that for cost analysis instead of 0.38 m diameter.

Height of the column (H) = $0.61 \text{ m} \cdot 20 \text{ stages} = 12.2 \text{ m}$

Reboiler duty calculations

Feed (F) conditions

Methanol vapor mole % = $[(45/32) / \{(45/32) + (55/18)\}] \cdot 100 = 31.3\%$

Saturated water vapor mole % = 68.7%

Temperature of the vapor = 92°C

$$H_{F, W} = 2663.3 \cdot 18 \cdot 0.687 = 32934.4 \text{ Kj/Kgmol} \quad [\text{using steam tables}]$$

$$H_{F, M} = \{81.6 \cdot 64.5 + 32256 + 44.06 \cdot (92 - 64.5)\} \cdot 0.313 = 12122.8 \text{ Kj/Kgmol}$$

$$\text{Now, } H_F = H_{F, W} + H_{F, M}$$

or,

$$H_F = 45057.2 \text{ Kj/Kgmol}$$

Distillate (D) conditions

$H_{ve} = 100\%$ methanol saturated vapors at 64.5°C

$$= 81.6 \times 64.5 + 32256 = 37520.0 \text{ Kj/Kgmol}$$

$$H_{le} = 5260 \text{ Kj/Kgmol} = H_D$$

Distillate (D) contains 0.09 mole % moisture.

Bottoms (B) conditions

$H_{ls} = 100\%$ saturated water at 100°C

$$= 419 \times 18 = 7542.0 \text{ Kj/Kgmol} = H_B$$

$H_{vs} =$ Saturated water vapor at 100°C

$$= 2676.0 \times 18 = 48168.0 \text{ Kj/Kgmol}$$

Bottoms (B) contains 0.28 mole % methanol.

Sizing calculations for the 5MGY biodiesel production plant

$$F = 674.0 \text{ Kg/hr} = 35.60 \text{ Kgmol/hr}$$

$$X_F = 0.313$$

$$X_B = 2.8 \times 10^{-3}$$

$$X_D = 0.999$$

From (13) and (14), we get,

$$D = 11.10 \text{ Kgmol/hr}$$

$$B = 24.5 \text{ Kgmol/hr}$$

$$L_c = 6 \times 11.10 = 66.70 \text{ Kgmol/hr}$$

$$V_c = 66.70 + 11.10 = 77.80 \text{ Kgmol/hr}$$

$$q_c = 77.80 \times 32256 = 2510807.1 \text{ Kj/hr (Condenser duty)}$$

After plugging the values in (15), we get

$$q_r = 1149140.0 \text{ Kj/hr}$$

Reboiler area calculations

$$U = 2250 \text{ W/m}^2/\text{K} \text{ (Peters et al., 2003)}$$

$$\Delta T_{\text{lmtd}} = (180-100) = 80.0^\circ\text{C}$$

From (16), we get,

$$A = 1.8 \text{ m}^2$$

With 15% oversize factor, we get,

$$A' = 2.1 \text{ m}^2$$

Sizing calculations for the 30MGY biodiesel production plant

$$F = 4031.0 \text{ Kg/hr} = 213.0 \text{ Kgmol/hr}$$

$$X_F = 0.313$$

$$X_B = 2.8 \times 10^{-3}$$

$$X_D = 0.999$$

Using (13) and (14), we get,

$$D/213.0 = \{(0.313) - (2.8 \times 10^{-3})\} / \{(0.999) - (2.8 \times 10^{-3})\}$$

$$D = 66.5 \text{ Kgmol/hr}$$

$$B = 146.6 \text{ Kgmol/hr}$$

$$L_e = 6 \times 66.5 = 399.0 \text{ Kgmol/hr}$$

$$V_e = 399.0 + 66.5 = 465.5 \text{ Kgmol/hr}$$

$$q_c = 465.5 \times 32256 = 15015910.0 \text{ Kj/hr (Condenser duty)}$$

From (15), $FH_F + q_r = DH_D + BH_B + q_c$, we get,

$$q_r = 6868185.0 \text{ Kj/hr}$$

Reboiler area calculations

$$U = 2250 \text{ W/m}^2/\text{K} \text{ (Peters et al., 2005)}$$

$$\Delta T_{\text{lmtd}} = (180-100) = 80.0^\circ\text{C}$$

From (16), $q_r = UA\Delta T_{\text{lmtd}}$, we get,

$$A = 10.6 \text{ m}^2$$

With 15% oversize factor, we get,

$$A' = 12.2 \text{ m}^2$$

(X) Condenser

Assumptions (Peters et al., 2005; Sheehan et al., 1998; Tapasvi et al., 2005)

1. Double pipe heat exchanger with carbon steel tube and shell construction.

Pressure – upto 4135 KPa

2. Condensing medium is incoming water at 20°C . The outgoing water is assumed to be at 35°C .

3. Temperature of stream methanol vapor and methanol liquid after condensing is 64.5°C

Sizing calculations

$$q_c = UA\Delta T_{\text{lmtd}} \tag{17}$$

where, q_c is obtained from the calculations performed for methanol recovery distillation column (W)

$$\Delta T_{\text{lmtd}} = [(64.5-20) - (64.5-35)] / [\ln\{(64.5-20)/(64.5-35)\}] = 36.56^\circ\text{C}$$

$$U = 1875 \text{ W/m}^2/\text{K} \text{ (Peters et al., 2003)}$$

For the 5 MGY biodiesel production plant

$$q_c = 2510807.1 \text{ Kj/hr}$$

Using (17), we get,

$$A = 10.2 \text{ m}^2$$

With 15% oversize factor, we get,

$$A' = 11.8 \text{ m}^2$$

For the 30 MGY biodiesel production plant

$$q_c = 15015910.0 \text{ Kj/hr}$$

Using (17), we get,

$$A = 60.95 \text{ m}^2$$

With 15% oversize factor, we get,

$$A' = 70.1 \text{ m}^2$$

Centrifugal pumps

Centrifugal pumps were assumed between equipment: (B) to (C), (C) to (E), NaOH solution to (E), (Soft water) to (E), (E) to (F), (H) to (I), (I) to (J), methanol to (J) and (L) and (K) to (L).

Assumptions (Peters et al., 2005; Zhang et al., 2003)

1. Inline vertical motor centrifugal pumps
2. The constant discharge pressure head of 1035 kPa for all the pumps
3. One spare pump
4. Minimum required volumetric flow rate and discharge head product is 3.0 for costing purposes

Sizing calculations

These pumps were sized based on the product of volumetric flow rates (m^3/s) and the discharge pressure head required (kPa), as summarized in Table 18.

Table 18. Centrifugal pump specifications for both plants

Centrifugal Pumps	Flow rate (m^3/s)* discharge pressure (KPa) [5 MGY]	Flow rate (m^3/s)* discharge pressure (kPa) [30 MGY]
1. B to C	3.0	4.6
2. C to E	3.0	4.5
3. NaOH to E	3.0	3.0
4. Soft water to E	3.0	3.0
5. E to F	3.0	5.2
6. H to I	3.0	4.4
7. I to J	3.0	4.4
8. Methanol to J and L	3.0	3.0
9. K to L	3.0	4.8
10. Spare	3.0	5.2

References

- Perry, R.H., and D.W. Green. 1997. *Perry's Chemical Engineers' Handbook*. New York, N.Y.: McGraw Hill.
- Peters, M.S., K.D. Timmerhaus, and R.E. West. 2003. *Plant design and economics for chemical engineering*. 5th edition. New York, N.Y.: McGraw Hill.
- Saravacos, G.D., and A.E. Kostaropoulos. 2002. *Handbook of Food Processing Equipment*. New York, N.Y.: Kluwer Academic/Plenum Publishers.
- 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.

Zhang, Y., M.A.Dube, D.D. McLean, and M. Kates. 2003. Biodiesel production from waste cooking oil: 1. Process design and technological assessment. *Bioresour. Technol.* 89 (1), 1-16.

APPENDIX B

BIODIESEL PROCESS MODEL SPREADSHEETS

This includes screen views of the various streams that were identified in the biodiesel process model spreadsheets along with the other information gathered for performing mass and energy balance calculations. The formulas used for the calculations have been displayed in the respective cells.

Mass balance spreadsheet

Stream	1	2	3	4	5
Descr.	Crude Oil	Heater Outstream	Phosphoric acid solution	Soft water	Mixing tank Outstream
Component					
Water (wt)			=D001*B33*0.15	=D.75*0.22	=E16+D16
Methanol (wt)					
Sodium Methoxide (wt)					
Crude Oil					
TG (g)	=B7.25	=B30			=C30
FFA (g)	=D.5	=B21			=C21
Phosphatides (g)	=E.25	=B22			D
Others (Unsuspendable matter) (g)	=1	=B23			=C23
Glycerol (g)					
Soaps (g)					
Methyl Ester (wt)					
NaOH (wt)					
Phosphoric acid (wt)			=D.001*B33*0.65		
HCl (wt)					
Gums (wt)					=C22+D28
Steam (wt)					
NaCl (wt)					
Total (M)	=SUM(B16:B32)	=SUM(C16:C32)	=SUM(D16:D32)	=SUM(E16:E32)	=SUM(F16:F32)
Temperature (C)	=D0	=D0			
Average specific heat (kJ/kg°C)	=B87	=B87			
Enthalpy (kJ/kg) (M)	=B25*B36	=C35*C36			
Mass Enthalpy (kW) (M/sec)	=B33*B37	=C33*C37			
Phase	Liquid	Liquid			

Stream	6	7	8	9	10
Descr.	Gums water mix	Gums	Water	De-gummed Oil	NaOH Solution
Component					
Water (wt)	=D.995*F16		=G16	=F16-G16	=I100.95*9.5*Q7
Methanol (wt)					
Sodium Methoxide (wt)					
Crude Oil					
TG (g)	=D.5/100*F.30	=G28		=F.30-G.30	
FFA (g)				=F.21	
Phosphatides (g)				=F.23	
Others (Unsuspendable matter) (g)					
Glycerol (g)					
Soaps (g)					
Methyl Ester (wt)					=J2.13*Q99
NaOH (wt)					
Phosphoric acid (wt)					
HCl (wt)					
Gums (wt)	=F30	=G30			
Steam (wt)					
NaCl (wt)					
Total (M)	=SUM(G16:G32)	=SUM(H16:H32)	=SUM(I16:I32)	=SUM(J16:J32)	=SUM(K16:K32)
Temperature (C)					
Average specific heat (kJ/kg°C)					
Enthalpy (kJ/kg) (M)	=J36*J36				=K35*K36
Mass Enthalpy (kW) (M/sec)	=J33*J37				=K33*K37
Phase	Liquid				Liquid

A		L	M
10 Stream		11	12
11 Descr.		Wash water	Refining tank Outstream
12			
13			
14 Component			
15			
16 Water (wt)		=15/100*M33	=L16+(B63:B60 *(99/100)*J21)+K*5*M16
17 Methanol (wt)			
18 Sodium Methoxide (wt)			
19 Crude Oil			
20 TG (g)			=J20
21 FFA (g)			=1/100*J21
22 Phosphatides (g)			=J22
23 Others (Unresponsible matter) (g)			=J23
24 Glycerol (g)			
25 Soaps (g)			=((B30:O:O:1)+O:O:23)*B30/((99+J21)*J21)
26 Methyl Ester (wt)			
27 NaOH (wt)			=((K*27/ B69:B60 *(99/100)*J21))
28 Phosphoric acid (wt)			
29 HCl (wt)			
30 Gums (wt)			
31 Steam (wt)			
32 NaCl (wt)			
33 Total (M)		=SUM(L16:L32)	=SUM(M16:M32)
34			
35 Temperature (C)		J35	J35
36 Average specific heat (kJ/kg°C)		=B91	=((M16+M27)*M33*4.184)+((M20+M21+M22+M23+M25)*M33*887)
37 Enthalpy (kJ/kg) (H)		=L35*L36	=M36*M36
38 Mass Enthalpy (MW) (kJ/sec)		=L37*L37	=M37*M37
39 Phase		Liquid	Liquid

A		N	O	P
10 Stream		13	14	15
11 Descr.		Waste	Centrifuge Outstream	Heater Outstream
12				
13				
14 Component				
15				
16 Water (wt)		=29.5/100*M16	=M16*N16	=O16
17 Methanol (wt)				
18 Sodium Methoxide (wt)				
19 Crude Oil				
20 TG (g)		=((J21+M21)*2.5	=M20*N20	=O20
21 FFA (g)			=M21	=O21
22 Phosphatides (g)			=M22	=O22
23 Others (Unresponsible matter) (g)		=O:O15*M20	=M23*N23	=O23
24 Glycerol (g)				
25 Soaps (g)		=M25		
26 Methyl Ester (wt)				
27 NaOH (wt)		=M27		
28 Phosphoric acid (wt)				
29 HCl (wt)				
30 Gums (wt)				
31 Steam (wt)				
32 NaCl (wt)				
33 Total (M)		=SUM(N16:N32)	=SUM(O16:O32)	=SUM(P16:P32)
34				
35 Temperature (C)			J35	J35
36 Average specific heat (kJ/kg°C)			=((O16/O:O33*B91)+((O20+O21+O22+O23)/O:O33*B87)	=O36
37 Enthalpy (kJ/kg) (H)			=O36*O36	=P36*P36
38 Mass Enthalpy (MW) (kJ/sec)			=O37*O37	=P37*P37
39 Phase			Liquid	Liquid

A	O	R	S	T
10 Stream	16	17	18	19
11 Descr.	Water vapor	Hot Oil	Refined Oil recovered 2	Catalyst in CSTR 1
12				
13				
14 Component				
15				
16 Water (w)	=C16			
17 Methanol (m)				= (T16-T18) * T18
18 Sodium Methoxide (sm)				= 1/100 * S20
19 Crude Oil				
20 TG (t)		=O20	=R20	
21 FFA (f)		=O21	=R21	
22 Phosphatides (p)		=O22	=R22	
23 Others (Unsuspendable matter) (o)		=O23	=R23	
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 (nc)				
33 Total (M)	=SUM(Q16:Q32)	=SUM(R16:R32)	=SUM(S16:S32)	=SUM(T16:T32)
34				
35 Temperature (C)	30	50	20	20
36 Average specific heat (kJ/kg°C)	1.94	=B87	=B87	=B89
37 Enthalpy (kJ/kg) (H)	=Q35*Q36	=R35*R36	=S35*S36	=T35*T36
38 Mass Enthalpy (kW) (M*H/sec)	=Q37*Q37	=R37*R37	=S37*S37	=T37*T37
39 Phase	Vapor	Liquid	Liquid	Liquid

A	U	V
10 Stream	20	21
11 Descr.	Methanol in CSTR 1	CSTR 1 Outstream
12		
13		
14 Component		
15		
16 Water (w)	=((B64*B83/B63)*S20)/T17	=((T17+U17)/((B64*B3)/B63)*(0.85*S20))-(B64*B80)*S21
17 Methanol (m)		=T19/(1.297/0.001)*S21
18 Sodium Methoxide (sm)		
19 Crude Oil		
20 TG (t)		=S20/(0.85*S20)
21 FFA (f)		0
22 Phosphatides (p)		=S22
23 Others (Unsuspendable matter) (o)		=S23
24 Glycerol (g)		=((B65/B63)*(0.85*S20))
25 Soaps (s)		=((B61)/500)*S21
26 Methyl Ester (me)		=((B66/B63)/(0.85*S20))
27 NaOH (n)		
28 Phosphoric acid (pa)		
29 HCl (h)		
30 Gums (gu)		
31 Steam (st)		
32 NaCl (nc)		
33 Total (M)	=SUM(U16:U32)	=SUM(V16:V32)
34		
35 Temperature (C)	20	50
36 Average specific heat (kJ/kg°C)	=B87	=((V17+V16)/(33*B89))-(V20+V21+V22+V23+V25+V26)/(33*B87)+(V24/V33*B90)
37 Enthalpy (kJ/kg) (H)	=U35*U36	=V35*V36
38 Mass Enthalpy (kW) (M*H/sec)	=U37*U37	=V37*V37
39 Phase	Liquid	Liquid

A		V		X		Y	
10	Stream	22		23		24	
11	Descr.		Glycerol phase 1		Fatty phase		Catalyst in CSTR 2
12							
13							
14	Component						
15							
16	Water (w)						
17	Methanol (m)		=V17*60/100		=V17-W17		=Y16*16-Y18
18	Sodium Methoxide (sm)		=V18				=1/100*Z20
19	Crude Oil						
20	TG (t)				=V20		
21	FFA (f)				=V21		
22	Phosphatides (p)				=V22		
23	Others (Unsaponifiable matter) (o)				=V23		
24	Glycerol (g)		=V24				
25	Soaps (s)		=10/100*V25		=V25-W25		
26	Methyl Ester (me)				=V26		
27	NaOH (n)						
28	Phosphoric acid (pa)						
29	HCl (h)						
30	Gums (gu)						
31	Steam (st)						
32	NaCl (na)						
33	Total (t)		=SUM(W16:W32)		=SUM(X16:X32)		=SUM(Y16:Y32)
34							
35	Temperature (C)						
36	Average specific heat (kJ/kg°C)						
37	Enthalpy (kJ/kg) (H)						
38	Mass Enthalpy (kW) (H*W/sec)						
39	Phase						
40							

A		Z		AA	
10	Stream	25		26	
11	Descr.		Methanol in CSTR 2		CSTR 7 Outstream
12					
13					
14	Component				
15					
16	Water (w)				
17	Methanol (m)		=(((B64*B65/B63)*QD)-(Y17+((X17+Y17*Z17)-(B64*B65)*QD-(1-(B4/100))*S20))))		
18	Sodium Methoxide (sm)				
19	Crude Oil				
20	TG (t)		=X20*(1-B4/100)		
21	FFA (f)		=X21		
22	Phosphatides (p)		=X22		
23	Others (Unsaponifiable matter) (o)		=X23		
24	Glycerol (g)		=((B65/B63)*QD-((1-(B4/100))*S20)))		
25	Soaps (s)		=X25		
26	Methyl Ester (me)		=((B66*B63)*QD-((1-(B4/100))*S20))*X26		
27	NaOH (n)				
28	Phosphoric acid (pa)				
29	HCl (h)				
30	Gums (gu)				
31	Steam (st)				
32	NaCl (na)				
33	Total (t)		=SUM(Z16:Z32)		=SUM(AA16:AA32)
34					
35	Temperature (C)				
36	Average specific heat (kJ/kg°C)				
37	Enthalpy (kJ/kg) (H)				
38	Mass Enthalpy (kW) (H*W/sec)				
39	Phase				
40					

A		AB	AC
10 Stream		27	28
11 Descr.		Glycerol phase 2	Ester phase
12			
13			
14 Component			
15			
16 Water (w)			
17 Methanol (m)		=AA17*60+100	=AA17*AB17
18 Sodium Methoxide (sm)		=Y18	
19 Crude Oil			
20 TG (t)			=AA20
21 FFA (f)			=AA21
22 Phosphatides (p)			=AA22
23 Others (Unsataponifiable matter) (o)			=AA23
24 Glycerol (g)		=AA24	
25 Soaps (s)		=10/100*AA25	=AA25*AB25
26 Methyl Ester (me)			=AA26
27 NaOH (n)			
28 Phosphoric acid (pa)			
29 HCl (h)			
30 Gums (gu)			
31 Steam (st)			
32 NaCl (na)			
33 Total (t)		=SUM(AB16:AB32)	=SUM(AC16:AC32)
34			
35 Temperature (C)			56
36 Average specific heat (kJ/kg°C)			= (AC17/AC33*889) + (AC20+AC21+AC22+AC23+AC25+AC26)/AC33*889
37 Enthalpy (kJ/kg) (H)			=AC36*AC38
38 Mass Enthalpy (kW) (H*W/sec)			=AC33*AC37
39 Phase			Liquid

A		AD	AE	AF	AG
10 Stream		29	30	31	32
11 Descr.		Heater	Wash water	Heater	Waste stream
12		Outstream		outstream	
13					
14 Component					
15					
16 Water (w)			=0.2*AC36	=AE16	=90/100*AF16
17 Methanol (m)		=AC17			=AC17
18 Sodium Methoxide (sm)					
19 Crude Oil					
20 TG (t)		=AC20			
21 FFA (f)		=AC21			
22 Phosphatides (p)		=AC22			
23 Others (Unsataponifiable matter) (o)		=AC23			
24 Glycerol (g)					
25 Soaps (s)		=AC25			=AC25
26 Methyl Ester (me)		=AC26			
27 NaOH (n)					
28 Phosphoric acid (pa)					
29 HCl (h)					
30 Gums (gu)					
31 Steam (st)					
32 NaCl (na)					
33 Total (t)		=SUM(AD16:AD32)	=SUM(AE16:AE32)	=SUM(AF16:AF32)	=SUM(AG16:AG32)
34					
35 Temperature (C)		70	70	70	
36 Average specific heat (kJ/kg°C)		=AC36	=B91	=B91	
37 Enthalpy (kJ/kg) (H)		=AD36*AD36	=AE35*AE36	=AF35*AF36	
38 Mass Enthalpy (kW) (H*W/sec)		=AD33*AD37	=AE33*AE37	=AF33*AF37	
39 Phase		Liquid	Liquid	Liquid	

A	AH	AI	AJ	AK
10 Stream	33 Washed esters	34 Aqueous phase	35 Esters	36 Heater Outstream
11 Descr.				
12				
13				
14 Component				
15				
16 Water (w)	=AF16/AJ16	=AH16-(0.5/100*AF16)	=AH16 - AJ16	=AJ16
17 Methanol (m)				
18 Sodium Methoxide (sm)				
19 Crude Oil				
20 TG (t)	=AJ20		=AJ20	=AJ20
21 FFA (f)	=AJ21		=AJ21	=AJ21
22 Phosphatides (p)	=AJ22		=AJ22	=AJ22
23 Others (Unresponisible matter) (o)	=AJ23		=AJ23	=AJ23
24 Glycerol (g)				
25 Soaps (s)				
26 Methyl Ester (me)	=AJ26		=AJ26	=AJ26
27 NaOH (n)				
28 Phosphoric acid (pa)				
29 HCl (h)				
30 Gums (gu)				
31 Steam (st)				
32 NaCl (na)				
33 Total (M)	=SUM(AH16:AH32)	=SUM(AI16:AI32)	=SUM(AJ16:AJ32)	=SUM(AK16:AK32)
34				
35 Temperature (C)			B0	95
36 Average specific heat (kJ/Kg°C)			=(AJ16/AJ33*B01)+(AJ26/AJ33*B07)	=AJ06
37 Enthalpy (kJ/Kg) (H)			=AJ35*AJ36	=AJ35*AJ36
38 Mass Enthalpy (kW) (M*H/sec)			=AJ33*AJ37	=AJ33*AJ37
39 Phase			Liquid	Liquid

A	AL	AM
10 Stream	37 Water vapor	38 Methyl esters (Biodiesel)
11 Descr.		
12		
13		
14 Component		
15		
16 Water (w)	=AJ16	
17 Methanol (m)		
18 Sodium Methoxide (sm)		
19 Crude Oil		
20 TG (t)		=AJ20
21 FFA (f)		=AJ21
22 Phosphatides (p)		=AJ22
23 Others (Unresponisible matter) (o)		=AJ23
24 Glycerol (g)		
25 Soaps (s)		
26 Methyl Ester (me)		=AJ26
27 NaOH (n)		
28 Phosphoric acid (pa)		
29 HCl (h)		
30 Gums (gu)		
31 Steam (st)		
32 NaCl (na)		
33 Total (M)	=SUM(AL16:AL32)	=SUM(AM16:AM32)
34		
35 Temperature (C)	B6	95
36 Average specific heat (kJ/Kg°C)	1.94	=B67
37 Enthalpy (kJ/Kg) (H)	=AL35*AL36	=AM35*AM36
38 Mass Enthalpy (kW) (M*H/sec)	=AL33*AL37	=AM33*AM37
39 Phase	Vapor	Liquid

A		AN
10 Stream		39
11 Descr.		Glycerol/Aqueous phase
12		
13		
14 Component		
15		
16 Water (w)		=AG16*AI15
17 Methanol (m)		=AN17*AB17+AG17
18 Sodium Methoxide (sm)		=VW13*AB13
19 Crude Oil		
20 TG (t)		
21 FFA (f)		
22 Phosphatides (p)		
23 Others (Unsaponifiable matter) (o)		
24 Glycerol (g)		=AN24+AB24
25 Soaps (s)		=VW25+AG25+AB25
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)		=SUM(AN16:AN32)
34		
35 Temperature (C)		20
36 Average specific heat (kJ/kg°C)		=(AN16/AN33*891)+(AN17/AN33*899)+(AN20/AN21+AN22+AN23+AN25/AN33*887)+(AN24/AN33*890)
37 Enthalpy (kJ/kg) (H)		=AN35*AN36
38 Mass Enthalpy (kW) (M*H/sec)		=AN33*AN37
39 Phase		Liquid

A		AO	AP
10 Stream		40	41
11 Descr.		Heater Outstream	Super heated Steam
12			
13			
14 Component			
15			
16 Water (w)		=AN16	
17 Methanol (m)		=AN17	
18 Sodium Methoxide (sm)		=AN18	
19 Crude Oil			
20 TG (t)			
21 FFA (f)			
22 Phosphatides (p)			
23 Others (Unsaponifiable matter) (o)			
24 Glycerol (g)		=AN24	
25 Soaps (s)		=AN25	
26 Methyl Ester (me)			
27 NaOH (n)			
28 Phosphoric acid (pa)			
29 HCl (h)			
30 Gums (gu)			
31 Steam (st)			=((AO16*891*(100-84.64/5)))+(AO16*851240)+(AO17*899*(4.5+AO35))+(AO17*EB841))EB835
32 NaCl (na)			
33 Total (M)		=SUM(AO16:AO32)	=SUM(AP16:AP32)
34			
35 Temperature (C)		245	250
36 Average specific heat (kJ/kg°C)		=AN36	
37 Enthalpy (kJ/kg) (H)		=AO35*AO36	2973.67
38 Mass Enthalpy (kW) (M*H/sec)		=AO33*AO37	=AP23*AP37
39 Phase		Liquid	Vapor
40			

A	AQ	AR	AS
10 Stream	42 Methanol Stream	43 Methanol Vapors	44 Bottoms
11 Unact.	Vapors		
12	Saturated		
13			
14 Component			
15			
16 Water (w)	=AQ16		=AQ16
17 Methanol (m)	=AQ17	=AQ17-(J 5/100*AQ17)	=AQ17-AR17
18 Sodium Methoxide (sm)			
19 Crude Oil			
20 TG (t)			
21 FFA (f)			
22 Phosphatides (p)			
23 Others (Unresponifiable matter) (o)			
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)	=AP31		
32 NaCl (na)			
33 Total (M)	=SUM(AQ16:AQ32)	=SUM(AR16:AR32)	=SUM(AS16:AS32)
34			
35 Temperature (C)	100.034		
36 Average specific heat (kJ/kg°C)			
37 Enthalpy (kJ/kg) (H)	=((B89*55)-(E1B41))* (E1B43)		
38 Mass*Enthalpy (MW) (H*H/sec)	=(((B89*55)*(E1B41)*AQ17) + ((AQ31+AQ16)*2575.1)/Vapor)		
39 Phase			
40			

A	AT	AU
10 Stream	45 Methanol	46 Hot glycerine solution
11 Unact.		
12		
13		
14 Component		
15		
16 Water (w)	=AR17	
17 Methanol (m)		=AQ18
18 Sodium Methoxide (sm)		
19 Crude Oil		
20 TG (t)		
21 FFA (f)		
22 Phosphatides (p)		
23 Others (Unresponifiable matter) (o)		
24 Glycerol (g)		=AQ24
25 Soaps (s)		=AQ25
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)	=SUM(AT16:AT32)	=SUM(AU16:AUS2)
34		
35 Temperature (C)		100.084
36 Average specific heat (kJ/kg°C)		=((AU18/AU33*B89)+(AU20+AU21+AU22+AU23+AU25/AU33*B87)+(AU24/AU33*B80))
37 Enthalpy (kJ/kg) (H)		=AU25*AU36
38 Mass*Enthalpy (MW) (H*H/sec)		=AU33*AU37
39 Phase		Liquid
40		

A	AV	AW	AX
10 Stream	47	48	49
11 Unact.	Glycerine	HCl	Acidulation reagent outstream
12			
13			
14 Component			
15			
16 Water (w)		=50/100*AV33-AV29	=AW16
17 Methanol (m)			=BE4/BS7*AV18
18 Sodium Methoxide (sm)	=AU18		
19 Crude Oil			
20 TG (t)			
21 FFA (f)			= (BB8/BB1*AV25)+AV21
22 Phosphatides (p)			
23 Others (Unreportable matter) (o)			
24 Glycerol (g)	=AU24		=AV24
25 Soaps (s)	=AU25		
26 Methyl Ester (me)			
27 NaOH (n)			
28 Phosphoric acid (pa)			
29 HCl (h)		=50/100*AV33*15/100	=AW29/(B71/BB1*AV25)/(B71/BS7*AV16)
30 Gums (gu)			
31 Steam (st)			
32 NaCl (na)			= (B73/BB1*AV25)+(BE4/BS7*AV16)
33 Total (M)	=SUM(AV16:AV32)	=SUM(AW16:AW32)	=SUM(AX16:AX32)
34			
35 Temperature (C)			
36 Average specific heat (kJ/Kg/C)			
37 Enthalpy (kJ/Kg) (H)			
38 Mass Enthalpy (kW) (M*H/sec)			
39 Phase			
40			

A	AY	AZ	BA	BB	BC
10 Stream	50	51			
11 Unact.	Waste (FFA/Unreacted Oil)	Glycerine outstream			
12					
13					
14 Component					
15					
16 Water (w)					
17 Methanol (m)		=AX17			
18 Sodium Methoxide (sm)					
19 Crude Oil					
20 TG (t)					
21 FFA (f)	=AX21				
22 Phosphatides (p)					
23 Others (Unreportable matter) (o)					
24 Glycerol (g)		=AX24			
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)	=SUM(AY16:AY32)	=SUM(AZ16:AZ32)			
34					
35 Temperature (C)					
36 Average specific heat (kJ/Kg/C)					
37 Enthalpy (kJ/Kg) (H)					
38 Mass Enthalpy (kW) (M*H/sec)					
39 Phase					
40					

A	B	C	D	E	F
41 Fatty acid composition of oils					
42					
43 Fatty acid	C 16	C 16:1	C 18	C 18:1	C 18:2
44 Triglyceride Molecular weight (gm)	807.339	801.288	891.501	885.453	879.405
45 Composition of the oil(% age)					
46 Soybean Oil	11	0	4	23	54
47 Canola Oil	4	25	2	80	20
48 Molecular weight fraction	11	0	4	23	54
49	=(B44*B45/100)	=(C44*C45/100)	=(D44*D45/100)	=(E44*E45/100)	=(F44*F45/100)
50 Molecular weight of the oil	=SUM(B48:F48)				
51					
52 Components of crude oil					
53	Soybean Oil	Canola Oil			
54					
55 Triglyceride	=(D 95*B6)	=(D 97*E6)			
56 Free fatty acids	=(D 100*B6)	=(D 100*E6)			
57 Phosphatides	=(D 107*B6)	=(D 112*E6)			
58 Others	=(D 112*B6)	=(D 111*E6)			
59					
61 Molecular weights					
62	g-mole				
63 Molecular weight of oil	=B60				
64 Molecular weight of methanol	=D34				
65 Molecular weight of glycerol	=D31				
66 Molecular weight of methyl esters	=(B63 + (D35*4) - B65)/3				
67 Molecular weight of sodium methoxide	54				
68 Molecular weight of water	18				
69 Molecular weight of NaOH	40				
70 Molecular weight of NaCl	=23+35.5				
71 Molecular weight of H ₂ O	=1+35.5				
72					

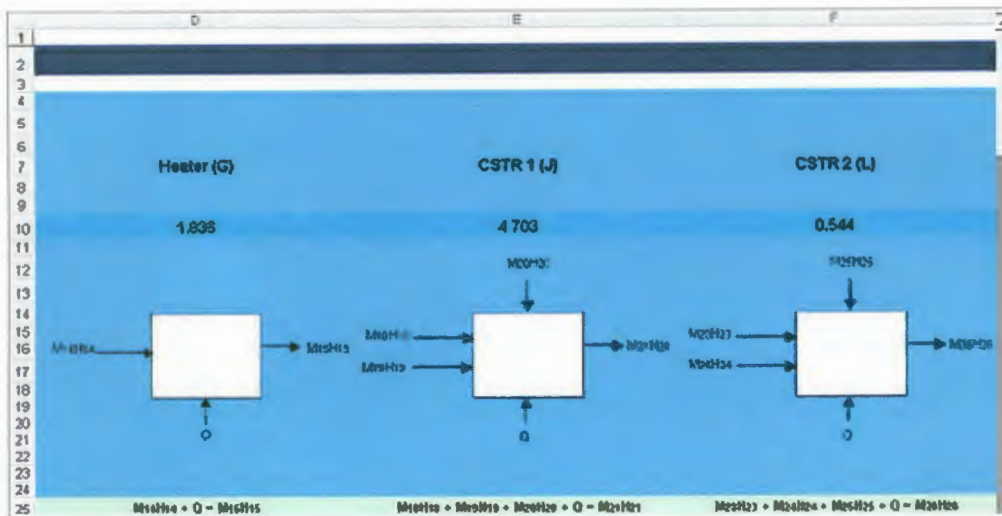
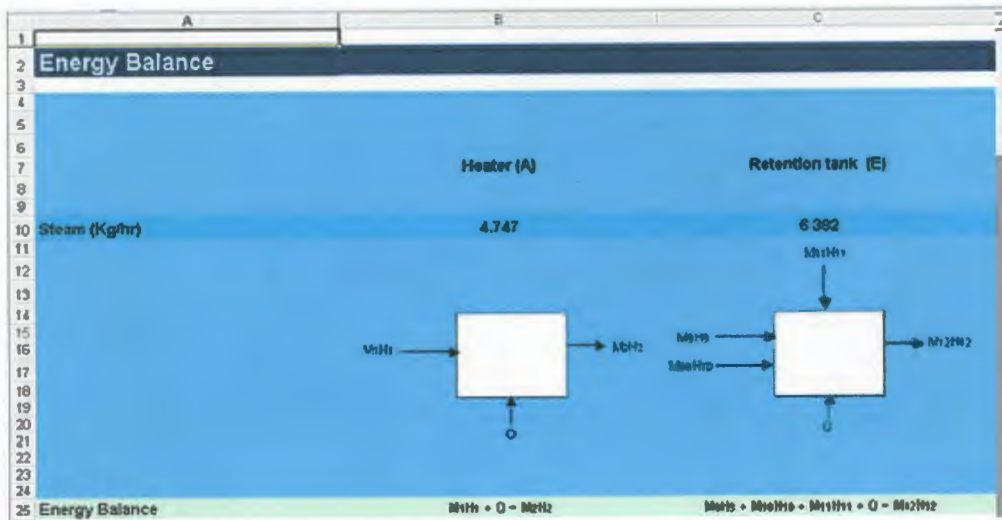
A	G	H	I	J	K
41 Fatty acid composition of oils					
42					
43 Fatty acid	C 18:3	C 20	C 22	C 22:1	C 24
44 Triglyceride Molecular weight (gm)	873.357	975.678	1059.878	1053.798	1113.906
45 Composition of the oil(% age)					
46 Soybean Oil	8	0	0	0	0
47 Canola Oil	10.5	1.5	0	1.5	0
48 Molecular weight fraction	8	0	0	0	0
49	=(G44*G45/100)	=(H44*H45/100)	=(I44*I45/100)	=(J44*J45/100)	=(K44*K45/100)
50 Molecular weight of the oil					

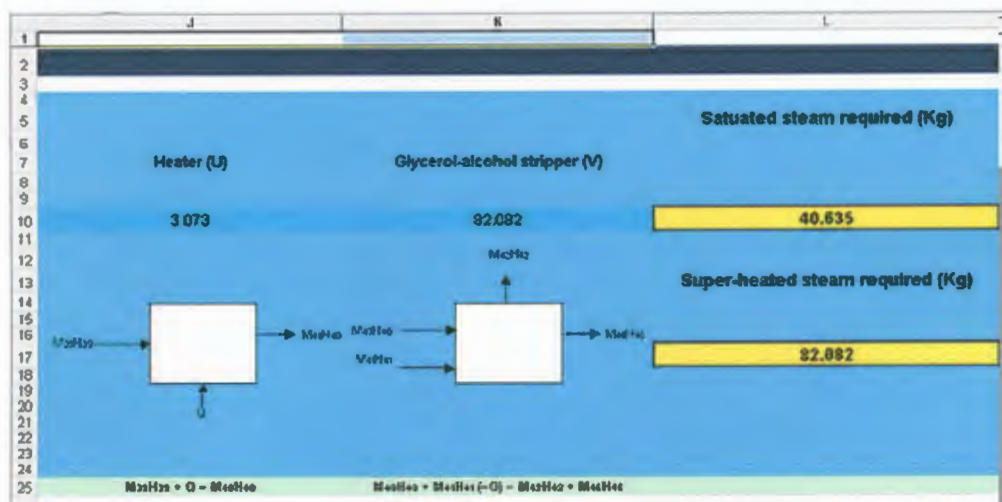
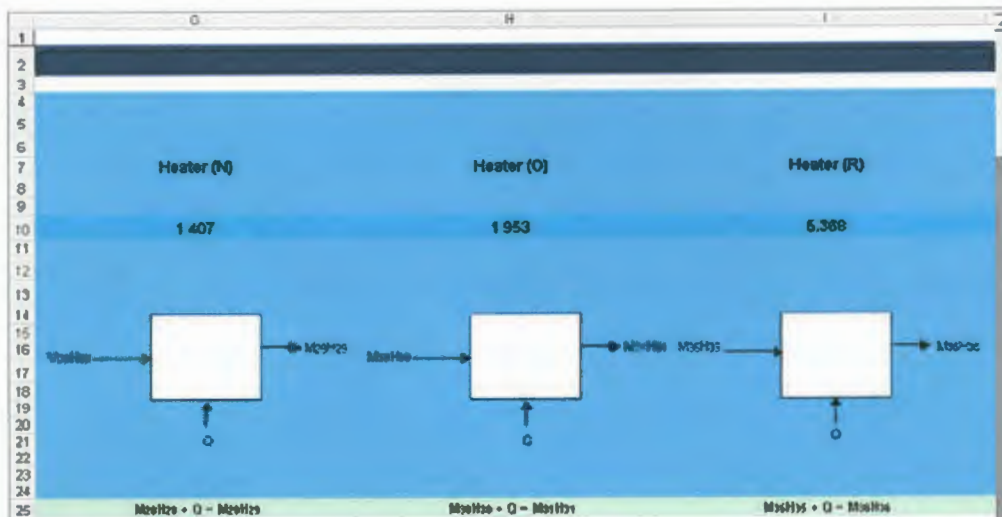
A	B	C	D	E	F
72					
73 Fatty acid	C 16	C 16:1	C 18	C 18:1	C 18:2
74 Molecular weight (g-mole)	286.4	284.4	284.4	282.4	280.4
75 Soybean oil free fatty acid MW (g-mole)	=(G74*(C45/100))	=(C74*(C45/100))	=(D74*(D45/100))	=(E74*(E45/100))	=(F74*(F45/100))
76 Canola oil free fatty acid MW (g-mole)	=(G74*(C47/100))	=(C74*(C47/100))	=(D74*(D47/100))	=(E74*(E47/100))	=(F74*(F47/100))
77					
78 Soy FFA MW (g-mole)	278.22				
79 Canola FFA MW (g-mole)	281.733				
80 Molecular weight of the FFA (g mole)	281.733				
81 Molecular weight of soap (g mole)	=(D80 + D81 + 0.02)				
82					
83 Specific heat	KJ/Kg/C				
84					
85 Soybean oil/ Soy biodiesel	1.917				
86 Canola Oil/Canola biodiesel	1.913				
87 Oil/Biodiesel	1.913				
88					
89 Methanol	2.55				
90 Glycerol	2.32				
91 Water	4.185				

A	G	H	I	J	K
72					
73 Fatty acid	C 18:3	C 20	C 22	C 22:1	C 24
74 Molecular weight (g-mole)	278.4	312.5	340.5	338.5	368.6
75 Soybean oil free fatty acid MW (g-mole)	=(G74*(C45/100))	=(H74*(C45/100))	=(I74*(I45/100))	=(J74*(J45/100))	=(K74*(K45/100))
76 Canola oil free fatty acid MW (g-mole)	=(G74*(C47/100))	=(H74*(C47/100))	=(I74*(I47/100))	=(J74*(J47/100))	=(K74*(K47/100))

A	B	C	D	E	F
93 Reactions Involved					
94					
95 FFA-NaOH (Alkali refining)	$RCOOH + NaOH$	\longrightarrow	$RCOONa + H_2O$		
96					
97 Amount of soap produced	$=((B81/B80)*(99/100)*J21)$				
98 Amount of water produced	$=((B68/B90)*(99/100)*J21)$				
99 Amount of sodium hydroxide consumed	$=((B69/B90)*(99/100)*J21)$	(99% FFA removal)	$=((B69/B80)*J21)$	(100% FFA removal)	
100					
101 FFA-Sodium methoxide (CSTR 1)	$RCOOH + NaOCH_3$	\longrightarrow	$RCOONa + CH_3OH$		
102					
103 Amount of soap produced	$=((B81/B80)*S21)$				
104 Amount of methanol produced in CSTR 1	$=((B64/B90)*S21)$				
105 Amount of sodium methoxide consumed	$=((B67/B90)*S21)$				
106					
107 Sodium methoxide-HCL	$NaOCH_3 + HCL$	\longrightarrow	$CH_3OH + NaCl$		
108 (Acidulation reactor)					
109 Amount of methanol produced	$=B64/B67*AV18$				
110 Amount of sodium chloride produced	$=B64/B67*AV18$				
111 Amount of hydrogen chloride consumed	$=B71/B67*AV18$				
112					
113 Soap-HCL (Acidulation reactor)	$RCOONa + HCL$	\longrightarrow	$RCOOH + NaCl$		
114					
115 Amount of FFA produced	$=B80/B81*AV25$				
116 Amount of sodium chloride produced	$=B70/B81*AV25$				
117 Amount of hydrogen chloride consumed	$=B71/B81*AV25$				

Energy balance spreadsheet





A	B	C
31	Evaporation energy of the saturated steam (KJ/Kg) [$h_g - h_f$]	2015.00
32	(Saturated steam at 180 C, 150 psi pressure)	
33	(Difference between steam and condensation)	
35	Available energy from the super-heated steam (KJ/Kg) [$h_{180} - h_g$]	717.43
36	(3 mbars gauge pressure, 250 C)	(Saturation temperature = 100.084 C)
39	Latent heat of vaporization	KJ/Kg
40	Water at 100.084 C	2256.44
41	Methanol at 64.5 C	1000.80
43	Methanol Distillation Column Reboiler duty	(KJ)
44	(Adapted from NREL report)	(Kcal)
45	(53.676 Kcal per 1000 kg biodiesel produced)	21403.04
46		5115.45