Biodiesel by supercritical transesterification

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BIODIESEL
BY SUPERCritical TRANSESTERIFICATION:
PROCESS DESIGN AND ECONOMIC FEASIBILITY

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2005
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1. Background

It is estimated that in the coming years, the fossil oil price will increase because the oil production can not meet the projected demand due to oil depletion (www.energycrisis.com). This is a result of overconsumption in the developed countries and overpopulation in the developing countries.

![Image of oil consumption and CO₂ emission between developed and developing countries]

*Figure 1. Oil consumption and CO₂ emission between developed and developing countries*

![Image of United States vs China comparison]

*Figure 2. Population, gas emission and oil consumption comparison between United States of America and China*
It can be seen from figure 1, that the oil demands and global warming gas emission of the developed countries is much higher than the ones of developing countries. Figure 2 shows us that the oil consumption and gas emission of United States of America is twice higher than Chinese although US population is only a quarter of China. This can be attributed to the higher oil consumption per person in the developed countries consequently leads to higher CO₂ emissions.

According to the Association for the Study of Peak Oil and Gas (ASPO) (http://en.wikipedia.org/wiki/Hubbert_peak#Peak_prediction), the world faces the start of oil depletion around 2007 and natural gas is expected to peak from 2010 to 2020 and then declines to inevitable zero as can be seen in the figure below.

It can be summarized that the current oil production can not meet the projected oil demand thus exploring new oil fields or developing an alternative fuel is essential.

A lot of efforts have been carried out to develop an alternative fuel for the current energy and transportation vehicle system, i.e.: fuel cell, electric power, hydrogen or natural gas for internal combustion engines, etc. One of the promising alternatives that is applied in small scale production is biodiesel.

The American Society for Testing and Materials (ASTM) defines biodiesel fuel as monoalkyl esters of long chain fatty acids derived from a renewable lipid feed stocks, such as vegetable oil or animal fat. "Bio" represents its renewable and biological source in contrast to traditional petroleum based diesel fuel; "diesel" refers to its use in diesel engines. As an alternative fuel, biodiesel can be used in neat form or mixed with petroleum based diesel.

Several sources for producing biodiesel have been studied such as rape seed, coal seed, palm oil, sunflower oil, waste cooking oil, soybean oil, etc. Due to the high cost of the fresh vegetable oil, waste cooking oil gave interesting properties because it can tackle environmental issue and it is available with relatively cheap price.
2. Theory

The most common way to produce biodiesel is by transesterification, which refers to a catalyzed chemical reaction involving vegetable oil and an alcohol to yield fatty acid alkyl esters (biodiesel) and glycerol (by-product) as can be seen in the figure below.

\[
\begin{align*}
&\text{Triacylglycerol} & \text{Methanol} & \text{Fatty acid methyl esters} & \text{Glycerol} \\
&(\text{Triglyceride}) & & (\text{FAME}) & \\
\end{align*}
\]

Figure 4. Transesterification reaction

Transesterification reactions can be alkali-catalyzed, acid-catalyzed or enzyme catalyzed. The first two types were extensively studied, because the enzyme catalyzed system requires a much longer reaction time than the other 2 systems (Watanabe et al, 2002). An excess of methanol is used to shift the reaction to the right side in order to achieve high yield of methyl esters / biodiesel.

Most biodiesel industries use alkali catalyzed process. One limitation to the alkali catalyzed process is its sensitivity to both water and free fatty acids. Free fatty acids can react with the alkali catalyst to produce soaps and water. Therefore Freedman et al (1984) stated that refined vegetable oils with free fatty acids content of less than 0.5% (acid value less than 1) should be used to maximize methyl esters formation.

The presence of water may cause ester saponification (Basu and Norris, 1996). Water can consume the catalyst and reduce the catalyst efficiency. The presence of water has a greater negative effect than that of the free fatty acids. Ma et al (1998) stated that the water content should be kept below 0.06%.

Most industries use pre-treatment step to reduce the free fatty acid and water content of the feed stream. Usually free fatty acid is reduced via an esterification reaction with methanol in the presence of sulfuric acid. The pre-treatment step not only causes the production process to be longer (Kusdiana and Saka, 2004) but also increase the capital cost.

These facts hinder the efficient use of waste cooking oil, animal fats and crude oils since they generally contain water and free fatty acids (Tomasevic and Marinkovic, 2003).

There is an alternative for biodiesel production, namely the supercritical methanol method. The great advantages of supercritical methanol are:
- no catalyst required
- not sensitive to both water and free fatty acid
- free fatty acids in the oil are esterified simultaneously
A comparison of the properties of the supercritical and conventional method can be seen below.

<table>
<thead>
<tr>
<th>Properties</th>
<th>Supercritical</th>
<th>Conventional</th>
</tr>
</thead>
<tbody>
<tr>
<td>Catalyst need</td>
<td>no (+)</td>
<td>yes</td>
</tr>
<tr>
<td>Reaction time</td>
<td>seconds - minutes</td>
<td>minutes - hours</td>
</tr>
<tr>
<td>Temperature (°C)</td>
<td>200-300</td>
<td>50-80</td>
</tr>
<tr>
<td>Pressure (bar)</td>
<td>100-200</td>
<td>1</td>
</tr>
<tr>
<td>Free fatty acid sensitive</td>
<td>no (+)</td>
<td>yes</td>
</tr>
<tr>
<td>Water sensitive</td>
<td>no (+)</td>
<td>yes</td>
</tr>
<tr>
<td>Pre-treatment</td>
<td>no (+)</td>
<td>yes</td>
</tr>
<tr>
<td>Catalyst removal</td>
<td>no (+)</td>
<td>yes</td>
</tr>
<tr>
<td>Soap removal</td>
<td>no (+)</td>
<td>yes</td>
</tr>
</tbody>
</table>

The absence of pre-treatment step, soap removal, and catalyst removal can significantly reduce the capital cost of a biodiesel plant, but the operating cost due to high temperature and pressure can be a drawback for supercritical method. That is why it is interesting to see whether the supercritical methanol method is economically feasible to be applied in a biodiesel plant.

3. Objectives
1. Develop a conceptual process design by noncatalytic transesterification from waste cooking oil
2. Carry out an economic feasibility study based on the process simulation
3. Carry out a sensitivity analysis

4. Process Design
4.1. Introduction

Complete process simulation were first carried out to assess the commercial feasibilities of the proposed processes. Most current simulation software can provide reliable information of the process operation because of their comprehensive thermodynamic packages, vast component libraries and advanced calculation techniques.

The process simulation software, Aspen Plus® version 11.1.1 developed by Aspen Technology Inc., Cambridge, Massachusetts USA was used in this research.

The procedures for process simulation mainly involve defining the chemical components, selecting a thermodynamic model, determining plant capacity, choosing proper operating units and setting up input conditions (flow rate, temperature, pressure and other conditions). Information on most components, such as methanol, glycerol, propane, and water is available in the Aspen Plus® component library.

Regarding the waste cooking oil or animal fat feedstock, oleic acid is considered as the major component of the oils and fats used in the food industry in the Netherlands and Europe (www.scientificpsychic.com). Triolein (C_{57}H_{106}O_{6}) was chosen to represent the waste cooking oil or animal fat in the Aspen Plus® simulation. Methyl oleate (C_{19}H_{36}O_{2}) was chosen as the fatty acid methyl ester as the biodiesel product.
The thermodynamic properties of triolein and methyl oleate, such as normal boiling point, specific gravity, standard enthalpy of formation, standard Gibbs energy of formation were modified to the values of oleic acid and methyl esters. The standard enthalpy of formation was calculated with the estimation method of Franklin. The standard Gibbs energy of formation was calculated with van Krevelen and Chermin method (Reid, Prausnitz and Sherwood, 1987) (further details can be seen in the appendix 5). Input of the realistic values of the thermodynamic properties to the simulation software was done with clicking the icon "component", then click icon "specification", and then click the "review" icon button. Then replace the thermodynamic data of triolein and methyl oleate with the ones of oleic acid and methyl esters.

Universal quasi-chemical (UNIQUAC) thermodynamic/ activity models were recommended due to the presence of highly polar components, i.e. methanol and glycerol. This is in close agreement with Zhanget al(2003). Some interaction parameter coefficients, such as for methanol/methyl oleate, glycerol/methyl oleate, propane/methyl oleate were estimated using the UNIQUAC liquid-liquid equilibrium module in Aspen Plus®.

The plant capacity was based on the data of Dutch Central Bureau of Statistics (www.cbs.nl). Based on the paid tax of waste cooking oil and animal fats value from the year 1999 to 2001, an average value per year of 96 million euro of waste oil / fat tax was covered. The tax for the waste cooking oil is 20 Euro cent / liter. Summarizing these 2 facts, a volume of 500 million liter of waste cooking oil and animal fat was calculated for the Netherlands. With the waste cooking oil density of 953 kg/m³, the available waste cooking oil in the Netherlands is 457,440 ton per year.

Some of the waste cooking oil and animal fat is recycled and used as a fertilizer, soap, and filler for cosmetics industry. That is why design capacities of 80,000 and 125,000 ton are considered realistic for the process design and simulation. The costs of these capacities will be compared with the values of conventional biodiesel plant and will be studied to determine the effect of capacities on the economic feasibility of the plant. Choosing proper operating units and setting up input conditions (flow rate, temperature, pressure and other conditions) will be explained in the next section.

4.2. Process Design and conditions

Waste cooking oil is preheated in heat exchanger B17 to 40 °C to decrease the viscosity and improve the flow property. Oil methanol mol ratio used in this process design is 1:24 and propane (propane methanol molar ratio 1:20) used as co-solvent.

Propane is chosen as a co-solvent because it was proven to decrease the supercritical temperature 320 to 280 °C, pressure from 400 to 128 bar, and methanol ratio from 42 to 24 (mol base), respectively (Cao et al, 2005). According to the author's experiment in pressurized autoclave, biodiesel yield was 98% in 10 minute reaction time.

Fresh streams of oil and make up of methanol are pumped to 5 bar pressure and mixed in B9 with the recycle stream of methanol and propane from the transesterification reaction and accommodated in the main reactor section (details can be seen in figure 6). Propane solvent is soluble at 40 °C and 5 bar which led to a reduction of a power consuming compressor.

A cascade of heat exchangers is used to integrate heat of the process. The transesterification reaction is carried out in a tubular reactor B1. Methanol and propane are recycled using 2 flash evaporators (B16, B8). Finally biodiesel and glycerol are obtained from settler unit (B11). The operating units will be explained further in the coming sections.
Conceptual Design of Noncatalytic transesterification:

Layer 1. Reactants and Products

- Waste cooking oil (15.9 ton/h)
- Methanol (1.8 ton/h)
- Biodiesel (16.0 ton/h)
- Glycerol (1.7 ton/h)

Layer 2. Process blocks

- Waste cooking oil (15.9 ton/h)
- Methanol (1.8 ton/h)
- Recycle 12 ton/h methanol 1 ton/h propane
- Biodiesel (16 ton/h)
- Glycerol (1.7 ton/h)

Layer 3. Process conditions

Figure 5. Conceptual design of noncatalytic transesterification
Figure 6. The Process Flow Diagram (PFD of the supercritical process simulation in Aspen Plus®)

Table 2. Input and result properties of the process simulation (125 kton/year capacity)

<table>
<thead>
<tr>
<th>Substream: MIXED</th>
<th>1</th>
<th>2</th>
<th>4</th>
<th>25</th>
<th>3</th>
<th>11</th>
<th>22</th>
<th>7</th>
<th>13</th>
<th>14</th>
<th>23</th>
<th>24</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mass Flow kg/hr</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>TRIOL-01</td>
<td>15938</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>METHA-01</td>
<td>0</td>
<td>1817</td>
<td>12023</td>
<td>13840</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>METHY-01</td>
<td>0</td>
<td>0</td>
<td>1</td>
<td>1</td>
<td>16012</td>
<td>16012</td>
<td>0</td>
<td>16012</td>
<td>1</td>
<td>16011</td>
<td>0</td>
<td>16011</td>
</tr>
<tr>
<td>GLYCE-01</td>
<td>0</td>
<td>0</td>
<td>34</td>
<td>34</td>
<td>1692</td>
<td>1692</td>
<td>0</td>
<td>1693</td>
<td>34</td>
<td>1659</td>
<td>1655</td>
<td>1</td>
</tr>
<tr>
<td>PROPA-01</td>
<td>0</td>
<td>0</td>
<td>952</td>
<td>952</td>
<td>952</td>
<td>734</td>
<td>219</td>
<td>218</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Total Flow kg/hr</td>
<td>15938</td>
<td>1817</td>
<td>13010</td>
<td>30765</td>
<td>30766</td>
<td>30766</td>
<td>2113</td>
<td>28649</td>
<td>10891</td>
<td>17759</td>
<td>1715</td>
<td>16040</td>
</tr>
<tr>
<td>Temperature K</td>
<td>298</td>
<td>298</td>
<td>313</td>
<td>553</td>
<td>512</td>
<td>393</td>
<td>379</td>
<td>379</td>
<td>316</td>
<td>316</td>
<td>298</td>
<td>298</td>
</tr>
<tr>
<td>Pressure atm</td>
<td>1</td>
<td>1</td>
<td>5</td>
<td>126</td>
<td>126</td>
<td>126</td>
<td>5</td>
<td>5</td>
<td>0</td>
<td>0</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td>Vapor Frac</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>1</td>
<td>0</td>
<td>1</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Liquid Frac</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>0</td>
<td>1</td>
<td>0</td>
<td>1</td>
<td>1</td>
<td>1</td>
</tr>
</tbody>
</table>
4.2.1. Transesterification

A heat exchanger B6 is used to change the temperature of the stream 17 first to 280 °C using hot stream 28. The transesterification reaction is carried out inside an adiabatic tubular reactor. The design of the tubular reactor is as the following.

Table 3. Design parameters of the supercritical transesterification reactor

<table>
<thead>
<tr>
<th>Properties</th>
<th>Unit</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Design Capacity</td>
<td>ton / year</td>
<td>125,000</td>
</tr>
<tr>
<td>Temperature</td>
<td>°C</td>
<td>280</td>
</tr>
<tr>
<td>Pressure</td>
<td>bar</td>
<td>128</td>
</tr>
<tr>
<td>Oil : methanol ratio</td>
<td>molar ratio</td>
<td>1 : 24</td>
</tr>
<tr>
<td>Propane : methanol ratio</td>
<td>molar ratio</td>
<td>1 : 20</td>
</tr>
<tr>
<td>Tubular Reactor</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Tube internal diameter</td>
<td>cm</td>
<td>10</td>
</tr>
<tr>
<td>Tube thickness</td>
<td>mm</td>
<td>7</td>
</tr>
<tr>
<td>Tube length</td>
<td>m</td>
<td>96</td>
</tr>
<tr>
<td>Number of tubes</td>
<td></td>
<td>21</td>
</tr>
<tr>
<td>Activation energy</td>
<td>kJ / kmol</td>
<td>38,482</td>
</tr>
<tr>
<td>Heat of reaction (slightly endothermic)</td>
<td>kJ / s</td>
<td>0.032</td>
</tr>
<tr>
<td>Reaction kinetics constant</td>
<td>second$^{-1}$</td>
<td>7 x 10$^{-3}$</td>
</tr>
<tr>
<td>Residence time</td>
<td>minutes</td>
<td>17</td>
</tr>
</tbody>
</table>

The reaction kinetics was based on the supercritical experimental works of Kusdiana and Saka (2001) and used as the parameters for the transesterification reaction.

The transesterification reaction is essential for the biodiesel production:

Triglyceride (Tg) + 3 Methanol (MOH) $\leftrightarrow$ 3 Methyl Esters (ME) + Glycerol (GI)

Tg refers to waste cooking oil. Three species can be defined as methyl esters (ME), glycerol (GI) and unmethyl esterified compounds (uME) which include triglyceride, diglyceride, monoglyceride and unreacted free fatty acids.

The reaction is assumed to proceed as a first order reaction, as a function of the decrease of molar flow of triglyceride and volume in a tubular plug flow reactor as the following

$$r_{ME} = -\frac{dF_{Tg}}{dV}$$  \hspace{1cm} (1)

Equation 1 can be modified to

$$r_{ME} = -\frac{dF_{uMe}}{dV}$$  \hspace{1cm} (2)
For a first order reaction, the rate of reaction can be written as the product of reaction rate \( k \) (s\(^{-1}\)) and the concentration of unconverted Methyl esters \([uME]\) as following

\[
r_{\text{ME}} = -k[uME]
\]

(3)

and can be modified to the following.

\[
k = \frac{\ln[uME_0] - \ln[uME]}{\theta}
\]

(4)

The reactor was modeled as a plug flow reactor (RPlug) in Aspen Plus with the properties as in the table 3. The transesterification reaction, activation energy and other kinetic parameters were entered as the reaction input. The activation energy was obtained by the calculation and estimation of the data of the Arrhenius plot of Kusdiana and Saka (2001)(appendix 4).

### Design of the tubular reactor B1

From the process simulation, an adiabatic tubular reactor is still sufficient to obtain high yield of biodiesel although the residence time increased from 9 minutes to 17 minutes. This can be attributed to the slightly endothermic reaction (table 3). But due to the significant decrease of operating cost to supply the constant heat inside the reactor (1 MW), adiabatic reactor is the best option to carry out transesterification.

Assuming that Reynolds number of 10,000 is sufficient to promote turbulent flow inside the reactor, the tube diameter and the length of the reactor was calculated by equation

\[
Re = \frac{\rho v d}{\eta}
\]

(5)

Re = dimensionless Reynolds number
\( \rho \) = mass density (kg/m\(^3\))
\( v \) = velocity (m/s)
\( d \) = diameter (m)
\( \eta \) = dynamic viscosity (kg/m.s)

The liquid dynamic viscosity \( 5 \times 10^{-4} \) kg/m.s was estimated by the method of van Velzen et al (Perry’s Chemical Engineer's Handbook page 2-365) (appendix 6). Reynolds number of 10,000 is chosen to calculate the correlation between the velocity and internal diameter,

\[
10,000 = \frac{315 \text{kg/m}^3 \cdot v \cdot d}{5 \times 10^{-4} \text{kg/m.s}} \quad \text{and} \quad v \cdot d = 1.73 \times 10^{-3} \text{m}^2/\text{s}
\]

in adiabatic reactor, the temperature went down to 240 °C with liquid density of 579 kg/m\(^3\). Based on the viscosity and temperature relation diagram (Sinnott, Coulson, 1998), Reynolds number of 12,141 – 14,839 was calculated (details in appendix 6). So, the turbulent flow still applies.
We know that the constant volumetric flow rate going in to the reactor is $2.74 \times 10^{-2} \text{ m}^3/\text{s}$, 

$$n_t \cdot \frac{1}{4} \cdot \pi \cdot d^2 \cdot v = 2.74 \times 10^{-2} \text{ m}^3/\text{s}$$

$$v \cdot d = 1.73 \times 10^{-2} \text{ m}^3/\text{s} \rightarrow v = \frac{1.73 \times 10^{-2}}{d} \text{ m}^3/\text{s}$$

$$n_t \cdot \frac{1}{4} \cdot \pi \cdot d^2 = \frac{2.74 \times 10^{-2}}{v} = \frac{2.74 \times 10^{-2} \cdot d}{1.73 \times 10^{-2}}$$

hence

$$d = \frac{1.6 \cdot \frac{4}{\pi} \cdot \frac{1}{n_t}}{v}, \text{ or } n_t = \frac{1.6 \cdot \frac{4}{\pi} \cdot \frac{1}{d}}{v}$$

So we know the relation between the number of tubes and internal diameter. The length of the reactor can be calculated by multiplying the velocity of the liquid with the residence time.

The proper internal diameter is chosen in relation to the tube thickness calculation due to the high temperature and pressure applied in the system.

To withstand the high temperature and pressure, minimum wall thickness is required. The following equations were used to calculate the design minimum tube thickness for the tubular reactor.

$$\sigma_{\text{max}} \leq \frac{S_{ut}}{3.5} \text{ (ASME 2001)} \quad (6)$$

ASME 2001 stands for American Society of Mechanical Engineers Pressure Vessel code 2001,

$$\sigma_{\text{max}} = \text{maximum allowable stress}$$

$$S_{ut} = \text{ultimate tensile strength}$$

$$S_{ut} = \text{for low carbon steels} = 2750 \text{ bar (ASTM SA-129)}$$

ASTM = American Society for Testing and Materials

$$\sigma_{\text{max}} \leq \frac{2750}{3.5} \rightarrow 916, \text{ at } 343 \degree C \text{ there is } 7.4\% \text{ increase of allowable stress,}$$

hence $$\sigma_{\text{max}} = 1.074 = 984,$$
According to Coulson & Richardson (1998),
The relation between wall thickness, diameter and design stress can be expressed:

For example, for internal diameter of 2.5 cm, the required minimum thickness is

\[ e = \frac{P_i (D_i + e)}{2 \sigma_{\text{max}}} \Rightarrow e = \frac{P_i \cdot D_i}{2 \sigma_{\text{max}} - P_i} = \frac{128 \cdot 0.025}{2 \cdot 984 - 128} = 0.0017 \text{ m} = 1.7 \text{ mm} \]

\( e \) = wall thickness
\( P_i \) = internal pressure
\( D_i \) = internal tube diameter

With the same calculation method described, the relation between internal diameter, tube thickness, number of tubes and length can be seen below.

<table>
<thead>
<tr>
<th>Internal diameter (cm)</th>
<th>Thickness of tube (mm)</th>
<th>( V ) (m/s)</th>
<th>Number of tubs ( N_t )</th>
<th>Tube length ( L_t ) (m)</th>
<th>Total volume of tubes (( m^3 ))</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0.70</td>
<td>1.73</td>
<td>202</td>
<td>954</td>
<td>4.2</td>
</tr>
<tr>
<td>2</td>
<td>1.39</td>
<td>0.87</td>
<td>101</td>
<td>477</td>
<td>4.2</td>
</tr>
<tr>
<td>3</td>
<td>2.09</td>
<td>0.57</td>
<td>68</td>
<td>315</td>
<td>4.2</td>
</tr>
<tr>
<td>4</td>
<td>2.78</td>
<td>0.43</td>
<td>51</td>
<td>236</td>
<td>4.2</td>
</tr>
<tr>
<td>5</td>
<td>3.48</td>
<td>0.34</td>
<td>41</td>
<td>188</td>
<td>4.2</td>
</tr>
<tr>
<td>6</td>
<td>4.17</td>
<td>0.29</td>
<td>34</td>
<td>157</td>
<td>4.2</td>
</tr>
<tr>
<td>7</td>
<td>4.87</td>
<td>0.25</td>
<td>29</td>
<td>136</td>
<td>4.2</td>
</tr>
<tr>
<td>8</td>
<td>5.57</td>
<td>0.21</td>
<td>26</td>
<td>116</td>
<td>4.2</td>
</tr>
<tr>
<td>9</td>
<td>6.26</td>
<td>0.19</td>
<td>23</td>
<td>103</td>
<td>4.2</td>
</tr>
<tr>
<td>10</td>
<td>6.96</td>
<td>0.17</td>
<td>21</td>
<td>92</td>
<td>4.2</td>
</tr>
<tr>
<td>11</td>
<td>7.65</td>
<td>0.15</td>
<td>19</td>
<td>84</td>
<td>4.2</td>
</tr>
<tr>
<td>12</td>
<td>8.35</td>
<td>0.14</td>
<td>17</td>
<td>79</td>
<td>4.2</td>
</tr>
<tr>
<td>13</td>
<td>9.04</td>
<td>0.13</td>
<td>16</td>
<td>71</td>
<td>4.2</td>
</tr>
<tr>
<td>14</td>
<td>9.74</td>
<td>0.12</td>
<td>15</td>
<td>66</td>
<td>4.2</td>
</tr>
<tr>
<td>15</td>
<td>10.43</td>
<td>0.11</td>
<td>14</td>
<td>61</td>
<td>4.2</td>
</tr>
<tr>
<td>16</td>
<td>11.13</td>
<td>0.11</td>
<td>13</td>
<td>58</td>
<td>4.2</td>
</tr>
<tr>
<td>17</td>
<td>11.83</td>
<td>0.10</td>
<td>12</td>
<td>56</td>
<td>4.2</td>
</tr>
<tr>
<td>18</td>
<td>12.52</td>
<td>0.10</td>
<td>11</td>
<td>54</td>
<td>4.2</td>
</tr>
<tr>
<td>19</td>
<td>13.22</td>
<td>0.09</td>
<td>11</td>
<td>49</td>
<td>4.2</td>
</tr>
<tr>
<td>20</td>
<td>13.91</td>
<td>0.09</td>
<td>10</td>
<td>48</td>
<td>4.2</td>
</tr>
</tbody>
</table>

Tubes having an internal diameter of 10 cm and thickness of 7 mm seem a good option to construct the reactor (assumed that maximum thickness of tube allowed 1 cm and maximum length of 100 m).
The 92 = 96 m length of the tube is constructed of 8 tubes of each 12 m length in series.
Then 21 units of these series in parallel construct the transesterification reactor.

The transesterification reaction is stopped immediately by cooling to 118 °C in a heat exchanger B15 with the help of stream 15 before enters the flash evaporator B16.
4.2.2. Methanol and propane recovery

The high pressure of the stream 11 was decreased to 5 bar inside a flash evaporator B16. Sudden decrease of the pressure accommodates the vapor phase of propane and methanol to occur, and flows to the upstream 22, which contains mainly methanol (89%) and propane (11%).

A vacuum flash evaporator B8 is needed to separate the remaining methanol and propane. The separation is carried out at pressure of 0.02 bar and temperature of 43 °C. These operating conditions are chosen because from sensitivity analysis results of the operating unit (appendix 7), it led to the optimum separation of methanol and propane mixture which still delivers biodiesel product with methanol content lower than maximum allowable by the European biodiesel standard EN 14214 (appendix 2). The capital cost and operating cost of a vacuum flash evaporator is considerably cheaper than vacuum distillation column. The upstream 13 contains 99% methanol and 1% propane.

The vacuum stream 13 is compressed to 5 bar before it enters mixer B4 with the stream 22. The high temperature of stream 21 is cooled immediately with the help of a heat exchanger B6 using cold stream 17 (the feed stream of oil, methanol and propane). The stream 20 is 94 °C so it needs to be cooled down with the help of cooling water stream 10 in heat exchanger B10. Stream 4 contains liquid methanol and propane mixture at 40 °C and 5 bar and then mixed with the fresh feed of oil and make up of methanol in mixer B9.

The total recovery of the methanol and propane is 99.3 %. The make up of methanol reacts with triglyceride to produce methyl esters and glycerol.

Design of the flash evaporator B16

The dimension of the flash evaporator tank was estimated by calculating the disengagement height, liquid level and diameter. A disengagement height equal to the vessel diameter should be provided above the liquid level. The liquid level will depend on the hold-up time for smooth operation and control, typically 10 minutes is allowed (Sinnott, Coulson and Richardson, 1998). But to decrease the material cost for the flash tank, 2 minutes is assumed to be a reasonable liquid hold-up time.

The maximum design vapor velocity can be calculated as the following

\[ u_v = 0.035 \sqrt{\frac{\rho_L}{\rho_v}} \]  

\( u_v \) = maximum design vapor velocity (m/s)  
\( \rho_L \) = liquid density (kg/m³)  
\( \rho_v \) = vapor density (kg/m³)
\[ u_\ast = 0.035 \sqrt{\frac{792}{5.09}} = 0.437 \text{ m/s} \]

Vapor volumetric flow rate = \( \frac{7987 \text{ kg/h}}{5.09 \text{ kg/m}^3 \cdot 3600 \text{ s}} = 0.436 \text{ m}^3/\text{s} \)

Vessel area = \( \frac{0.436 \text{ m}^3/\text{s}}{0.437 \text{ m/s}} = 0.99 \text{ m}^2 \)

Disengagement space = Diameter = \( \sqrt{\frac{4 \cdot 0.99}{\pi}} = 1.13 \text{ m} = 1.2 \text{ m} \)

Liquid volumetric flow = \( \frac{22764 \text{ kg/h}}{792 \text{ kg/m}^3 \cdot 3600 \text{ s}} = 7.98 \cdot 10^{-3} \text{ m}^3/\text{s} \)

Volume for 2 minutes hold up = \( 7.98 \cdot 10^{-3} \text{ m}^3/\text{s} \times 120 \text{ s} = 0.96 \text{ m}^3 \)

Liquid level/depth = \( \frac{0.96}{0.99} = 0.97 \text{ m} \)

Assuming that 30 cm is available for adjusting and implementing the inlet tube/pipie, the total height of the flash tank is 2.5 m and diameter of 1.2 m.

---

Figure 7. Flash evaporator tank design
Vacuum flash evaporator tank B8 dimension

With the same method, the dimension of the vacuum flash evaporator can be estimated;

\[ u_r = 0.035 \sqrt[2]{\frac{864}{2.48.10^{-2}}} = 6.53 \text{ m/s}, \]

Vapor volumetric flow rate = \( \frac{5006 \text{ kg/h}}{2.48.10^{-2} \text{ kg/m}^3 \cdot 3600 \text{ s}} = 56 \text{ m}^3/\text{s} \)

Vessel area = \( \frac{56 \text{ m}^3/\text{s}}{6.53 \text{ m/s}} = 8.6 \text{ m}^2 \)

Disengagement space = Diameter = \( \sqrt{\frac{4 \times 8.6}{\pi}} = 3.3 \text{ m} \)

Liquid volumetric flow = \( \frac{17758 \text{ kg/h}}{864 \text{ kg/m}^3 \cdot 3600 \text{ s}} = 5.7 \times 10^{-3} \text{ m}^3/\text{s} \)

Volume for 10 minutes hold up = \( 5.7 \times 10^{-3} \text{ m}^3/\text{s} \times 600 \text{ s} = 3.4 \text{ m}^3 \)

Liquid level/depth = \( \frac{3.4}{8.6} = 0.4 \text{ m} \)

The diameter of the vacuum flash tank is 3.3 m and height of 4 m as can be seen below.

![Figure 8. Vacuum flash evaporator tank design](image)
4.2.3. Glycerol separation

Stream 5 (17.7 ton/h) which contains mainly biodiesel as the end-product and glycerol as by-product needs to be cooled down to 25 °C using 17.7 ton / h cooling water in a heat exchanger B13. The outlet of the cooling water stream is used further to cool the methanol-propane recycle stream, the stream leaving the reactor, and to preheat the cold feed stream of waste cooking oil. The stream is cooled down with the help of a cooling tower (B20) 5.1 m in base diameter and 7.6 m in height. Then it is pumped back to heat exchanger B13.

Glycerol is separated using a settling tank as also very often and usually applied in the real biodiesel industries. Stream 23 contains pure glycerol (96.4%) as a by-product and stream 24 contains high purity biodiesel (99.8%) and passes the European biodiesel standard EN 14214 as can be seen below.

<table>
<thead>
<tr>
<th>Component</th>
<th>End Stream 24</th>
<th>European Biodiesel standard</th>
</tr>
</thead>
<tbody>
<tr>
<td>Biodiesel</td>
<td>99.82</td>
<td>96.50</td>
</tr>
<tr>
<td>Methanol</td>
<td>0.17</td>
<td>0.20 (maximum)</td>
</tr>
<tr>
<td>Glycerol</td>
<td>0.01</td>
<td>0.20 (maximum)</td>
</tr>
<tr>
<td>Triglyceride</td>
<td>0.00</td>
<td>0.20 (maximum)</td>
</tr>
<tr>
<td>Propane</td>
<td>0.00</td>
<td>-</td>
</tr>
</tbody>
</table>

Decanter design

Decanter / settler tank is designed using several parameters, i.e.: settling velocity of the droplets, droplet diameter, area of interface, droplet size and velocity of dispersed phase.

The decanter vessel is sized on the basis that the velocity of the continuous phase must be less than the settling velocity of the droplets of the dispersed phase as the following,

\[ v_c = \frac{L_c}{A_i} v_d \]  

where,

- \( v_c \) = velocity of the continuous phase (m/s)
- \( v_d \) = settling velocity of the dispersed phase droplets with diameter \( d_d \) (m/s)
- \( L_c \) = continuous phase volumetric flow rate (m/s)
- \( A_i \) = area of interface (m²)

Assuming that glycerol is the dispersed phase and 5 minutes residence time of the droplets in the dispersion band is satisfactory for the process (Coulson & Richardson, 1998)

<table>
<thead>
<tr>
<th>Component</th>
<th>Flow rate (kg/h)</th>
<th>Density (kg/m³)</th>
<th>Viscosity (Ns/m²)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Glycerol</td>
<td>1700</td>
<td>1261</td>
<td>0.95 x 10⁻³</td>
</tr>
<tr>
<td>Methyl esters</td>
<td>16000</td>
<td>900</td>
<td>3 x 10⁻³</td>
</tr>
</tbody>
</table>
Take assumption the droplet diameter \( (d_d) \) of 150 \( \mu m \).

The settling velocity can be calculated as the following,

\[
v_d = \frac{d_d^2 \cdot g \cdot (\rho_c - \rho_d)}{18 \cdot \mu_c}
\]

where,

- \( d_d \) = droplet diameter (m)
- \( v_d \) = settling velocity of the dispersed phase droplets with diameter \( d_d \) (m/s)
- \( \rho_c \) = density of the continuous phase (kg/m\(^3\))
- \( \rho_d \) = density of the dispersed phase (kg/m\(^3\))
- \( \mu_c \) = viscosity of the continuous phase (Ns/m\(^2\))
- \( g \) = gravitational acceleration = 9.81 m/s\(^2\)

The settling velocity of the glycerol droplets can be calculated,

\[
v_d = \frac{(150 \times 10^{-6})^2 \cdot 9.81 \cdot (900 - 1261)}{18 \cdot 0.95 \times 10^{-3}} = 1.5 \times 10^{-3} \text{ m/s}
\]

Using a vertical cylindrical vessel with the area of interface,

\[
A_i = \pi r^2, \quad L_c = \frac{16000 \text{ kg/h}}{900 \text{ kg/m}^3} = 4.9 \times 10^{-3} \text{ m}^3/\text{s} \quad \text{and} \quad A_i = \frac{4.9 \times 10^{-3} \text{ m}^3/\text{s}}{1.5 \times 10^{-3} \text{ m/s}} = 3.4 \text{ m}^2
\]

\[
r = \frac{3.4}{\pi} = 1 \text{ m}, \quad \text{diameter of vessel} = 2 \text{ m}
\]

Take the height of vessel is twice diameter, height = 4 m

Assume that the dispersion band is 10% of the height = 0.4 m,

the residence time of the droplets in the dispersion band is

\[
\frac{0.4}{v_d} = 0.4 \times 1.5 \times 10^{-3} = 280 \text{ s (4.7 min)}
\]

This is satisfactory, a time of 2 to 5 min is recommended (Coulson & Richardson, 1998).

Piping calculation:

Flow rate:

\[
\text{Flow rate} = \left( \frac{16000 \text{ kg/h}}{900 \text{ kg/m}^3} + \frac{1700 \text{ kg/h}}{1261 \text{ kg/m}^3} \right) x \frac{1}{3600} = 5.3 \times 10^{-3} \text{ m}^3/\text{s}
\]

Area of pipe:

\[
\text{Area of pipe} = \frac{5.3 \times 10^{-3} \text{ m}^3/\text{s}}{1 \text{ m}} = 5.3 \times 10^{-3} \text{ m}^2
\]

Pipe diameter:

\[
\text{Pipe diameter} = \sqrt{\frac{5.3 \times 10^{-3} \times 4}{\pi}} = 0.082 \text{ m} = 85 \text{ mm}
\]

Take the position of the interface as half-way up the vessel and

the light liquid off-take as at 90% of the vessel height, then

\[
z_1 = 0.9 \times 4 \text{ m} = 3.6 \text{ m}, \quad z_2 = 0.5 \times 4 \text{ m} = 2 \text{ m}
\]

\[
z_3 = \frac{(3.6 - 2) \times 900 + 2}{1261} = 3.1 \text{ m}
\]
The proposed design of decanter can be seen below.

![Decanter vessel](image)

**Figure 9. Decanter vessel**

### 4.3. Unit operations

To conclude the process design aspect, the summary of the main equipments and their utilities are described in the table below.

<table>
<thead>
<tr>
<th>Main process equipment</th>
<th>Block name</th>
<th>Equipment description</th>
<th>Process I 125,000 ton / year</th>
<th>Process II 80,000 ton / year</th>
<th>Process III 8,000 ton / year</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Transesterification reactor</strong></td>
<td>B1</td>
<td>Plug flow reactor</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>Temperature (°C)</td>
<td>280</td>
<td>280</td>
<td>280</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Pressure (bar)</td>
<td>128</td>
<td>128</td>
<td>128</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Number of tubes</td>
<td>21</td>
<td>34</td>
<td>4</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Tube diameter (cm)</td>
<td>10</td>
<td>6</td>
<td>10</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Tube length (m)</td>
<td>96</td>
<td>99</td>
<td>32</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Tube thickness (mm)</td>
<td>7</td>
<td>5</td>
<td>7</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Heat duty (kW)</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td><strong>Separation unit</strong></td>
<td>B16</td>
<td>Flash evaporator</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>Temperature (°C)</td>
<td>118</td>
<td>118</td>
<td>118</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Pressure (bar)</td>
<td>5</td>
<td>5</td>
<td>5</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Size (D x H) (m)</td>
<td>1.2 x 2.5</td>
<td>0.9 x 2.2</td>
<td>0.3 x 1.5</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Heat duty (kW)</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td></td>
<td>B8</td>
<td>Distillation column</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>Pressure (bar)</td>
<td>1.01</td>
<td>1.01</td>
<td>1.01</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Size (D x H) (m)</td>
<td>4.9 x 5.4</td>
<td>3.9 x 4.2</td>
<td>1.2 x 1.6</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Reboiler heat duty (kW)</td>
<td>5165</td>
<td>1255</td>
<td>128</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Condenser heat duty (kW)</td>
<td>-108</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>B11</td>
<td>Settler tank</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>Temperature (°C)</td>
<td>25</td>
<td>25</td>
<td>25</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Pressure (bar)</td>
<td>1</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Size (D x H) (m)</td>
<td>2 x 4</td>
<td>1.6 x 3.2</td>
<td>0.1 x 0.3</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Area of Interface (m²)</td>
<td>3.4</td>
<td>2</td>
<td>0.2</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Heat duty (kW)</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Main process equipment</td>
<td>Block name</td>
<td>Equipment description</td>
<td>Process I 125,000 ton / year</td>
<td>Process II 80,000 ton / year</td>
<td>Process II 8,000 ton / year</td>
</tr>
<tr>
<td>------------------------</td>
<td>-----------</td>
<td>-----------------------</td>
<td>-----------------------------</td>
<td>-----------------------------</td>
<td>-----------------------------</td>
</tr>
<tr>
<td></td>
<td>B6</td>
<td>Required area (m²)</td>
<td>1059</td>
<td>655</td>
<td>93</td>
</tr>
<tr>
<td></td>
<td>B15</td>
<td>Required area (m²)</td>
<td>109</td>
<td>69</td>
<td>7</td>
</tr>
<tr>
<td></td>
<td>B10</td>
<td>Required area (m²)</td>
<td>100</td>
<td>60</td>
<td>10</td>
</tr>
<tr>
<td></td>
<td>B13</td>
<td>Required area (m²)</td>
<td>31</td>
<td>0.1</td>
<td>2</td>
</tr>
<tr>
<td>Heat exchangers</td>
<td>B20</td>
<td>Cooling tower</td>
<td>Natural-draft 5.1 x 7.6</td>
<td>Natural-draft 3.2 x 4.9</td>
<td>Natural-draft 0.3 x 0.5</td>
</tr>
<tr>
<td></td>
<td>B14</td>
<td>Compressor work required (kW)</td>
<td>816</td>
<td>1849</td>
<td>189</td>
</tr>
<tr>
<td></td>
<td>B5</td>
<td>Pump work duty (kW)</td>
<td>284</td>
<td>195</td>
<td>39</td>
</tr>
<tr>
<td></td>
<td>B2</td>
<td>Pump work duty (kW)</td>
<td>10</td>
<td>8</td>
<td>0.6</td>
</tr>
<tr>
<td></td>
<td>B3</td>
<td>Pump work duty (kW)</td>
<td>1</td>
<td>1</td>
<td>0.1</td>
</tr>
<tr>
<td></td>
<td>B12</td>
<td>Pump work duty (kW)</td>
<td>1</td>
<td>0.5</td>
<td>0.05</td>
</tr>
<tr>
<td></td>
<td>B19</td>
<td>Pump work duty (kW)</td>
<td>1</td>
<td>1</td>
<td>0.02</td>
</tr>
</tbody>
</table>

5. Economic feasibility evaluation

After the process flow diagram completed in Aspen Plus, it was exported to Aspen Icarus Process Evaluator (IPE) to evaluate the capital cost of the operating units and operating costs for the default operating time of 8000 hours/ year. The currency used in the Aspen IPE is United States dollar.

Variables in the cost calculation are: plant location, production capacity, bio-ethanol scenario. The plant locations were simulated in the average United States location and in the Netherlands. These variables were studied for the sensitivity analysis. It is interesting to know the feasibility of using bio-ethanol instead methanol (which derived from natural gas, emits CO(2) in the process). Environmental friendly bio-ethanol was not possible to simulate in Aspen Plus software technically. The software does not have component of ethyl oleate or similar component which can represent ethyl esters (biodiesel produced from transesterification of oleic acid and ethanol).

Warabi et al (2004) studied the reactivity of triglycerides and fatty acids of rapeseed oil in supercritical alcohols. Using supercritical ethanol, 98% yield of ethyl esters was obtained in 45 minutes instead of 15 minutes using supercritical methanol (at their lab experiments conditions).

Assuming that the process condition is similar to the ones of methanol, only residence time is longer (factor of 3), the bio-ethanol cost calculation was performed using multiplication of factor 3 of the purchased reactor cost.
5.1. The capital cost calculation

The equipments cost which contributes to the capital cost was calculated from the data of DACE price book (DACE stands for Dutch Association of Cost Engineers) edition November 2003. The cost was corrected with the CEPCI ratio (CEPCI stands for Chemical Engineering's Plant Cost Index) as can be seen below.

<table>
<thead>
<tr>
<th>Year</th>
<th>CEPCI</th>
</tr>
</thead>
<tbody>
<tr>
<td>1998</td>
<td>389.5</td>
</tr>
<tr>
<td>2003</td>
<td>402.0</td>
</tr>
<tr>
<td>2004</td>
<td>444.2</td>
</tr>
<tr>
<td>March 2005</td>
<td>468.3</td>
</tr>
</tbody>
</table>

(Source: Chemical Engineering. Economic Indicators. July 2005)

Main equipments, i.e. compressor which is not included in the DACE price book were estimated from the price data of process design course given in Eindhoven University of Technology by former ABB Lummus director (Prof. F.M. Dautzenberg). Conversion of euro to United States dollar was also included in the cost calculation (appendix 8).

The purchase cost of equipments for 125,000 ton / year production capacity was calculated as follows:

**Reactor:**
Reactor construction: 21 parallel series
1 series contains of 8 tubes (internal diameter of 10cm, 7 mm thick) of 12 m length in series with 7 bends, then 21 of these series stack together to construct the reactor.
100 m of 4 inch diameter cost 1,210 euro, so 1,210 x 12/100 x 21 x 8 x (468.3/402) = 28,416.7 euro.

Isolation for the reactor:
From above, the series which contains 8 tubes in series = 8 x (10 cm+ 0.7 cm+0.7 cm) = 91.2 cm = 0.912 m wide and 12 m length = 0.9 x 12 m = 10.8 m² x 83 euro (isolatiedikte 100 voor vlakke wanden) = 896.4 euro x 2 = 1792.8 euro
From the side view, 12 m x 0.114 m= 1.368 x 83 euro = 113.544 x 2 = 227.088 euro
From the other side view, 0.912 m x 0.114 m=0.104 x 83 euro = 8.63 x 2 = 17.26 euro
Isolation = (1792.8 + 227.09 + 17.26) x (468.3/402) = 2373.13 euro x 21 = 49,835.73 euro
Total reactor purchase cost = 78,252.43 euro x 1.232 = 96,407 US $.

**Flash evaporator** with 1.13 m diameter and 2.39 m height; from DACE costbook a steel column with diameter of 1.5 m and height of 5 m costs 34,000 Euro = times the CEPCI index and time the conversion euro to dollar gives 78,074 US $.

**Vacuum flash evaporator** with 4.9 m diameter and 5.4 m height; steel tank of 10m height and 4 m diameter costs 85,000. Assumed this is similar, the purchase cost is 243,982 US $.

With the same method, the cost calculation was performed for the process equipments and the well known six-tenth rule (Sinnot, Coulson, 1998) is used to calculate the capital cost of the other capacities (80,000 and 8,000 ton/year) (details in the appendix 10).
The results can be seen in the table below.

<table>
<thead>
<tr>
<th>Plant capacity (ton/year)</th>
<th>125,000 bioethanol</th>
<th>125,000 bioethanol</th>
<th>80,000 bioethanol</th>
<th>80,000 bioethanol</th>
<th>8,000 bioethanol</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reactor</td>
<td>B1</td>
<td>97,884</td>
<td>293,652</td>
<td>74,889</td>
<td>224,668</td>
</tr>
<tr>
<td>Flash evaporator</td>
<td>B16</td>
<td>78,074</td>
<td>78,074</td>
<td>59,733</td>
<td>59,733</td>
</tr>
<tr>
<td>Distillation column</td>
<td>B8</td>
<td>36,763</td>
<td>36,763</td>
<td>28,127</td>
<td>28,127</td>
</tr>
<tr>
<td>Settler Tank</td>
<td>B11</td>
<td>53,102</td>
<td>53,102</td>
<td>40,627</td>
<td>40,627</td>
</tr>
<tr>
<td>HeatX 1</td>
<td>B6</td>
<td>398,204</td>
<td>398,204</td>
<td>304,659</td>
<td>304,659</td>
</tr>
<tr>
<td>HeatX 2</td>
<td>B10</td>
<td>59,445</td>
<td>59,445</td>
<td>45,481</td>
<td>45,481</td>
</tr>
<tr>
<td>HeatX 3</td>
<td>B15</td>
<td>54,537</td>
<td>54,537</td>
<td>41,725</td>
<td>41,725</td>
</tr>
<tr>
<td>HeatX 4</td>
<td>B17</td>
<td>34,445</td>
<td>34,445</td>
<td>26,353</td>
<td>26,353</td>
</tr>
<tr>
<td>Cooling tower</td>
<td>B20</td>
<td>118,272</td>
<td>118,272</td>
<td>90,488</td>
<td>90,488</td>
</tr>
<tr>
<td>Compressor</td>
<td>B14</td>
<td>709,527</td>
<td>709,527</td>
<td>542,846</td>
<td>542,846</td>
</tr>
<tr>
<td>Pump 1</td>
<td>B3</td>
<td>58,842</td>
<td>58,842</td>
<td>45,019</td>
<td>45,019</td>
</tr>
<tr>
<td>Pump 2</td>
<td>B5</td>
<td>13,060</td>
<td>13,060</td>
<td>9,992</td>
<td>9,992</td>
</tr>
<tr>
<td>Pump 3</td>
<td>B18</td>
<td>6,785</td>
<td>6,785</td>
<td>5,191</td>
<td>5,191</td>
</tr>
<tr>
<td>Pump 4</td>
<td>B12</td>
<td>6,785</td>
<td>6,785</td>
<td>5,191</td>
<td>5,191</td>
</tr>
<tr>
<td>Pump 5</td>
<td>B19</td>
<td>6,785</td>
<td>6,785</td>
<td>5,191</td>
<td>5,191</td>
</tr>
</tbody>
</table>

| EQUIPMENTS COST          | 1,732,510          | 1,928,278          | 1,325,512         | 1,475,291         | 332,954         | 370,576         |
| ISBL                     | 8,662,549          | 9,641,389          | 6,627,560         | 7,376,453         | 1,664,768       | 1,852,881       |
| OSBL                     | 1,732,510          | 1,928,278          | 1,325,512         | 1,475,291         | 332,954         | 370,576         |
| Fixed Capital            | 10,395,058         | 11,569,666         | 7,953,072         | 8,851,744         | 1,997,721       | 2,223,457       |

5.2. Operating cost calculation

Fixed capital for equipment cost or inside battery limit (ISBL) is the cost for processing units, i.e. reactors, mixers, heat exchangers, pumps, compressors, etc. This cost was calculated by multiplying the purchased total equipment cost with a factor of 5, so

Fixed capital for equipment cost or ISBL = purchased cost * 5

This factor is in agreement with 4.7 Lang factor for fluids processing plant (Sinnott, Coulson and Richardson, 1998).

The cost calculation method also includes Outside Battery Limit (OSBL) which covers tankage, yards, roads, and other general facilities. The normal default value is 20% of ISBL.

The fixed capital = ISBL + OSBL

The total plant capital cost = fixed capital + working capital + start-up cost
Working capital is the fund required for routine operation, including inventories, accounts receivable and payable and cash on hand. The result of the calculation can be seen below.

Table 10. Annual cost and required selling price of biodiesel by supercritical transesterification

<table>
<thead>
<tr>
<th>BIODIESEL BY SUPERCRITICAL TRANSESTERIFICATION 125,000 ton / year</th>
</tr>
</thead>
<tbody>
<tr>
<td>TOTAL CAPITAL COST</td>
</tr>
<tr>
<td>Fixed Capital                                               US $ 23,064,236</td>
</tr>
<tr>
<td>Start Up Cost                                               US $ 1,738,984</td>
</tr>
<tr>
<td>US $ 5,216,953</td>
</tr>
<tr>
<td>location                                                   United States</td>
</tr>
<tr>
<td>Plant Capacity                                             125,000 MT / year</td>
</tr>
<tr>
<td>VARIABLE COST                                              cost/unit</td>
</tr>
<tr>
<td>Raw Material                                               Waste cooking oil</td>
</tr>
<tr>
<td>Methanol</td>
</tr>
<tr>
<td>Total Raw Materials</td>
</tr>
<tr>
<td>Start Up / Recycling afterwards                             Methanol</td>
</tr>
<tr>
<td>Propane</td>
</tr>
<tr>
<td>Total Start-up</td>
</tr>
<tr>
<td>Utilities                                                  Electricity</td>
</tr>
<tr>
<td>Cooling water</td>
</tr>
<tr>
<td>Total Utilities</td>
</tr>
<tr>
<td>By-product Credit                                          Glycerol</td>
</tr>
<tr>
<td>Total By-product Credit</td>
</tr>
<tr>
<td>Fixed Cost                                                 Operating Labor</td>
</tr>
<tr>
<td>1 supervisors shift each 300 k$/year</td>
</tr>
<tr>
<td>Maintenance</td>
</tr>
<tr>
<td>Plant Overhead</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
</tr>
<tr>
<td>Total Fixed Cost</td>
</tr>
<tr>
<td>TOTAL OPERATING COST                                       20,867,811</td>
</tr>
<tr>
<td>REQUIRED SELLING PRICE                                      RSP (US $ / TON)</td>
</tr>
<tr>
<td>RSP (US $ / KG)</td>
</tr>
<tr>
<td>RSP (US $ / liter)</td>
</tr>
</tbody>
</table>
The operating cost calculation for the other capacities can be seen in the appendix 12. And the summary of the cost calculation for the biodiesel production in the United States can be seen below.

Table 11. The required selling price of biodiesel by supercritical transesterification in US

<table>
<thead>
<tr>
<th>Plant Capacity / Process Condition</th>
<th>125,000 ton/year</th>
<th>125,000 ton/year biodiesel</th>
<th>80,000 ton/year</th>
<th>80,000 ton/year biodiesel</th>
<th>8,000 ton/year</th>
<th>8,000 ton/year biodiesel</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fixed Capital</td>
<td>10,395,058</td>
<td>11,569,666</td>
<td>7,953,072</td>
<td>8,851,744</td>
<td>1,997,721</td>
<td>2,223,457</td>
</tr>
<tr>
<td>Working Capital</td>
<td>1,661,348</td>
<td>1,743,555</td>
<td>1,513,014</td>
<td>1,567,628</td>
<td>313,729</td>
<td>319,398</td>
</tr>
<tr>
<td>Start Up Cost</td>
<td>4,984,045</td>
<td>5,230,664</td>
<td>4,539,042</td>
<td>4,702,884</td>
<td>941,187</td>
<td>958,193</td>
</tr>
<tr>
<td>Total Capital Cost</td>
<td>17,040,452</td>
<td>18,543,885</td>
<td>14,005,128</td>
<td>15,122,255</td>
<td>3,252,638</td>
<td>3,501,048</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>location</th>
<th>United States</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>Annual VARIABLE COST</th>
</tr>
</thead>
<tbody>
<tr>
<td>Raw Material</td>
</tr>
<tr>
<td>Waste cooking oil</td>
</tr>
<tr>
<td>Methanol</td>
</tr>
<tr>
<td>Bioethanol</td>
</tr>
<tr>
<td>Total raw material cost</td>
</tr>
<tr>
<td>Start Up</td>
</tr>
<tr>
<td>Methanol / bio-ethanol</td>
</tr>
<tr>
<td>Propane</td>
</tr>
<tr>
<td>Total Start Up Cost</td>
</tr>
<tr>
<td>Utilities</td>
</tr>
<tr>
<td>Electricity</td>
</tr>
<tr>
<td>Cooling water</td>
</tr>
<tr>
<td>Biodiesel for the reboiler</td>
</tr>
<tr>
<td>Total Utilities Cost</td>
</tr>
<tr>
<td>By-product Credit</td>
</tr>
<tr>
<td>Glycerol</td>
</tr>
<tr>
<td>Total By-product Credit</td>
</tr>
<tr>
<td>Fixed Cost</td>
</tr>
<tr>
<td>Operating Labor</td>
</tr>
<tr>
<td>Maintenance</td>
</tr>
<tr>
<td>Plant Overhead</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
</tr>
<tr>
<td>Total Fixed Cost</td>
</tr>
<tr>
<td>TOTAL OPERATING COST</td>
</tr>
<tr>
<td>Capital Charges</td>
</tr>
<tr>
<td>S. G &amp; A</td>
</tr>
<tr>
<td>REQUIRED SELLING PRICE</td>
</tr>
<tr>
<td>RSP (US $ / TON)</td>
</tr>
<tr>
<td>RSP (US $ / KG)</td>
</tr>
<tr>
<td>RSP (US $ / liter)</td>
</tr>
</tbody>
</table>

P.S.: 20% return of investment (ROI) used
Working capital = 2 * (number of months of inventory) * (monthly operating cost) + (cost of initial load of catalyst)

In this case catalyst cost is 0, no catalyst used.
For biodiesel, which is large bulk chemicals number of months of inventory is 0.5.

Start-up cost represents "last minute" plant modifications, salaries and expenses of the start-up team and frequently, inefficient operations of the plant at the beginning.

Start-up cost = 1-4 months of operating costs

For "first of a kind" technology start-up cost = 3 months of operating costs
For "known" technology start-up cost = 1 month of operating costs

3 months was chosen for this study, because supercritical transesterification is considered a novel approach to utilize the waste cooking oil and to eliminate the need of pre-treatment step, but it is not 4 months because the transesterification reaction itself is known technology in the biodiesel industry.

The operating cost includes the cost of raw materials, start-up cost, utilities (electricity, cooling water) and fixed costs (operating labor, maintenance, plant overhead and taxes and insurance).

Since the plant complex is considered as isolated system, the tax is not income taxes, it is considered as the property taxes and other local taxes, and average is 2% of fixed capital. This is in agreement with Coulson & Richardson (1998).

The capital charges cost includes working capital, startup cost, annually fixed cost and fixed capital investment (ISBL+OSBL).

Taken into account for the capital charges:
15 years project life (excluding construction).
2 years construction time
50% income tax (46% federal plus 4% state tax), in agreement with Zhang et al (2003)
10% investment tax credit
20% return rate of investment (ROI)

For 2 years construction and 20% ROI,
The annual capital charges = 0.326*Fixed capital + 0.368*Working capital + 0.216*Start-up cost + 0.106*Fixed cost

S,G & A stands for sales, general and administrative cost which represents expenses, research and development (R&D), administrative cost beyond plant level and corporate overhead. It is 5% of the selling price (total operating cost and capital charges).

The required selling price (RSP) is the price of the product which is required to cover all costs (variable, fixed and overhead), recover the total investment and provide the specified return of the employed capital. Assumed that the density of the biodiesel (methyl esters) is 840 kg/m³, the RSP in 15 years for plant capacity of 125,000 ton/year is 0.20 US$/liter, for 80,000 ton/year is 0.27
US$ / liter, and 0.60 US$/liter for 8,000 ton/year. Using bioethanol for the reaction input, the prices are slightly more expensive, 0.21 US$ / liter (125,000 ton/year), 0.28 US$ / liter (80,000 ton/year), and 0.61 US$ / liter (8,000 ton/year).

Required selling price is proper criterion in the context of R&D work (Dautzenberg, 2003). But net present value or worth of the project is crucial for go-no go decision.

5.3. Net present value calculation

Net present worth / value is closely related to cash flow of project. In calculating cash flows, the project is usually considered as an isolated system, so the effect of depreciation of the investment and inflation are not considered (Sinnot, Coulson and Richardson, 1998). The net present value / worth can be calculated as follow

Net Present Value or Worth (NPV or NPW)

\[
NPV = \frac{\text{Estimated net cash flow in year } n \times (1 + r)^{-n}}{\text{(1 + r)}}
\]

where \( r \) is the discount rate / interest rate (percent/100)

\( \text{NFW} = \text{Net Future Worth} \)

and

\[
\sum_{t=1}^{t} \frac{\text{NFW}}{(1 + r)^n}
\]

\( t = \text{life of project, years} \)

Table 12. Net Present Value of the biodiesel plant (125 kton/year)

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0</td>
<td>-12</td>
<td>0</td>
<td>-11.6</td>
<td>-11.6</td>
<td>-9.7</td>
</tr>
<tr>
<td>2</td>
<td>0</td>
<td>-12</td>
<td>0</td>
<td>-11.6</td>
<td>-23.2</td>
<td>-8.0</td>
</tr>
<tr>
<td>3</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>-14.3</td>
<td>5.1</td>
</tr>
<tr>
<td>4</td>
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<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>-5.5</td>
<td>4.3</td>
</tr>
<tr>
<td>5</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>3.4</td>
<td>3.6</td>
</tr>
<tr>
<td>6</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>12.3</td>
<td>3.0</td>
</tr>
<tr>
<td>7</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>21.1</td>
<td>2.5</td>
</tr>
<tr>
<td>8</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>30.0</td>
<td>2.1</td>
</tr>
<tr>
<td>9</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>38.8</td>
<td>1.7</td>
</tr>
<tr>
<td>10</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>47.7</td>
<td>1.4</td>
</tr>
<tr>
<td>11</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>56.5</td>
<td>1.2</td>
</tr>
<tr>
<td>12</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>65.4</td>
<td>1.0</td>
</tr>
<tr>
<td>13</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>74.3</td>
<td>0.8</td>
</tr>
<tr>
<td>14</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>83.1</td>
<td>0.7</td>
</tr>
<tr>
<td>15</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>92.0</td>
<td>0.6</td>
</tr>
<tr>
<td>16</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>100.8</td>
<td>0.5</td>
</tr>
<tr>
<td>17</td>
<td>29.7</td>
<td>0</td>
<td>-20.9</td>
<td>8.9</td>
<td>109.7</td>
<td>0.4</td>
</tr>
</tbody>
</table>

Net Present Value of project 11.1
From table 12, for 15 years project life (excluding 2 years of construction), 20% return of investment (ROI) and 20% discount / interest rate, The NPV for the project (125 kton/year) is 11.1 million US $, 9.1 million US $ (80 kton / year) and 2.8 million US$ (8 kton/year) (20% ROI; 3.0 million US$ for 15% ROI) with payback time 5 years for both capacities.

It can be concluded that biodiesel by supercritical transesterification is technically and economically feasible. Comparison to other works can be seen below.

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Process type</td>
<td>125,000</td>
<td>80,000</td>
<td>100,000</td>
</tr>
<tr>
<td>Raw material</td>
<td>supercritical</td>
<td>supercritical</td>
<td>alkali catalysed</td>
</tr>
<tr>
<td></td>
<td>waste cooking oil</td>
<td>waste cooking oil</td>
<td>beef tallow</td>
</tr>
<tr>
<td>Total capital cost (US$)</td>
<td>23.2 million</td>
<td>18.5 million</td>
<td>12 million</td>
</tr>
<tr>
<td>Total operating cost (US$)</td>
<td>29.7 million</td>
<td>25.4 million</td>
<td>34 million</td>
</tr>
<tr>
<td>Biodiesel break even price (US$/ton)</td>
<td>352</td>
<td>411</td>
<td>340</td>
</tr>
<tr>
<td>Glycerol credit (US$)</td>
<td>15.9 million</td>
<td>6.2 million</td>
<td>6 million</td>
</tr>
<tr>
<td>1200 US$/ton (92% pure)</td>
<td>11.1 million</td>
<td>9.1 million</td>
<td>unknown</td>
</tr>
<tr>
<td>NPV</td>
<td>15 years</td>
<td>15 years</td>
<td>unknown</td>
</tr>
<tr>
<td>Project life</td>
<td>20</td>
<td>20</td>
<td>unknown</td>
</tr>
<tr>
<td>Return of investment (%)</td>
<td>20</td>
<td>20</td>
<td>16</td>
</tr>
</tbody>
</table>

Table 13. Cost calculation of biodiesel plants

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Process type</td>
<td>8,000</td>
<td>8,000</td>
</tr>
<tr>
<td>Raw material</td>
<td>supercritical</td>
<td>supercritical</td>
</tr>
<tr>
<td></td>
<td>waste cooking oil</td>
<td>waste cooking oil</td>
</tr>
<tr>
<td>Total capital cost (US$)</td>
<td>4.4 million</td>
<td>2.7 million</td>
</tr>
<tr>
<td>Total operating cost (US$)</td>
<td>3.8 million</td>
<td>7.1 million</td>
</tr>
<tr>
<td>Biodiesel break even price (US$/ton)</td>
<td>712</td>
<td>884</td>
</tr>
<tr>
<td>Glycerol credit (US$)</td>
<td>1 million</td>
<td>0.7 million</td>
</tr>
<tr>
<td>1200 US$/ton (92% pure)</td>
<td>2.8 million</td>
<td>unknown</td>
</tr>
<tr>
<td>NPV</td>
<td>15 years</td>
<td>unknown</td>
</tr>
<tr>
<td>Project life</td>
<td>20</td>
<td>15</td>
</tr>
<tr>
<td>Return of investment (%)</td>
<td>20</td>
<td>15</td>
</tr>
</tbody>
</table>

Table 14. Cost calculation of small biodiesel plants

From tables above, it can be concluded that supercritical bio-ethanol scenario is economically feasible. The increase of the reactor purchase cost to accommodate the use of bio-ethanol hardly influences the required selling price of the biodiesel produced. It differs only 1 US$ cent per liter.
5.4. The Dutch Scenario

A lot of waste cooking oil from the Netherlands is exported to Germany for biodiesel industries feedstock. It is interesting to know whether it is feasible to run biodiesel plant in the Netherlands. Assumed that the waste cooking oil is available for the price of 0.30 Euro / liter (Novem, 2003), with the same cost calculation the required selling price of biodiesel produced in the Netherlands can be seen below.

Table 15. Dutch required selling price of biodiesel by supercritical transesterification

<table>
<thead>
<tr>
<th>Plant Capacity (ton/year)</th>
<th>125,000</th>
<th>80,000</th>
<th>8,000</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fixed Capital</td>
<td>16,218,193</td>
<td>12,408,247</td>
<td>3,116,811</td>
</tr>
<tr>
<td>Working Capital</td>
<td>3,582,825</td>
<td>2,707,373</td>
<td>437,358</td>
</tr>
<tr>
<td>Start Up Cost</td>
<td>10,748,475</td>
<td>8,122,119</td>
<td>1,312,074</td>
</tr>
<tr>
<td>Total Capital Cost</td>
<td>30,549,493</td>
<td>23,237,740</td>
<td>4,866,242</td>
</tr>
<tr>
<td>Location</td>
<td>Netherlands</td>
<td>Netherlands</td>
<td>Netherlands</td>
</tr>
<tr>
<td>Annual VARIABLE COST</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Raw Material</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Waste cooking oil</td>
<td>48,175,498</td>
<td>31,016,925</td>
<td>3,158,471</td>
</tr>
<tr>
<td>Methanol</td>
<td>4,218,750</td>
<td>2,736,000</td>
<td>278,400</td>
</tr>
<tr>
<td>Bioethanol</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Total raw material cost</td>
<td>52,394,248</td>
<td>33,752,925</td>
<td>3,436,871</td>
</tr>
<tr>
<td>Start Up</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Methanol/bio-ethanol</td>
<td>14,400</td>
<td>9,050</td>
<td>924</td>
</tr>
<tr>
<td>Propane</td>
<td>4,409</td>
<td>2,672</td>
<td>269</td>
</tr>
<tr>
<td>Total Start Up Cost</td>
<td>18,809</td>
<td>11,722</td>
<td>1,193</td>
</tr>
<tr>
<td>Utilities</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Electricity</td>
<td>3,299,760</td>
<td>2,080,803</td>
<td>226,416</td>
</tr>
<tr>
<td>Cooling water</td>
<td>150,000</td>
<td>98,000</td>
<td>8,217</td>
</tr>
<tr>
<td>Total Utilities Cost</td>
<td>3,449,760</td>
<td>2,178,803</td>
<td>234,633</td>
</tr>
<tr>
<td>By-product Credit</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Glycerol</td>
<td>15,937,500</td>
<td>6,234,000</td>
<td>1,017,600</td>
</tr>
<tr>
<td>Total By-product Credit</td>
<td>15,937,500</td>
<td>6,234,000</td>
<td>1,017,600</td>
</tr>
<tr>
<td>Fixed Cost</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Operating Labor</td>
<td>1,020,000</td>
<td>1,020,000</td>
<td>1,020,000</td>
</tr>
<tr>
<td>Maintenance</td>
<td>756,849</td>
<td>579,052</td>
<td>579,052</td>
</tr>
<tr>
<td>Plant Overhead</td>
<td>967,370</td>
<td>931,810</td>
<td>931,810</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
<td>324,364</td>
<td>248,165</td>
<td>62,336</td>
</tr>
<tr>
<td>Total Fixed Cost</td>
<td>3,068,583</td>
<td>2,779,027</td>
<td>2,593,198</td>
</tr>
<tr>
<td>TOTAL OPERATING COST</td>
<td>42,993,901</td>
<td>32,488,478</td>
<td>5,248,294</td>
</tr>
<tr>
<td>Capital Charges</td>
<td>9,381,533</td>
<td>7,187,822</td>
<td>1,751,060</td>
</tr>
<tr>
<td>S, G &amp; A</td>
<td>2,618,772</td>
<td>1,983,815</td>
<td>349,968</td>
</tr>
<tr>
<td>REQUIRED SELLING PRICE</td>
<td>54,994,205</td>
<td>41,660,115</td>
<td>7,349,322</td>
</tr>
<tr>
<td>RSP (US $ / TON)</td>
<td>440</td>
<td>521</td>
<td>919</td>
</tr>
<tr>
<td>RSP (US $ / KG)</td>
<td>0.44</td>
<td>0.52</td>
<td>0.92</td>
</tr>
<tr>
<td>RSP (US $ / liter)</td>
<td>0.37</td>
<td>0.44</td>
<td>0.77</td>
</tr>
<tr>
<td>RSP (EURO / liter)</td>
<td>0.30</td>
<td>0.36</td>
<td>0.63</td>
</tr>
</tbody>
</table>
From the table 15, it can be concluded that the biodiesel plant is economically feasible in the Netherlands as it can be sold within the range of 0.30 – 0.63 Euro / liter. This is acceptable since Rice (1998) estimated that the biodiesel produced from waste cooking oil with conventional method could be sold with break even price of 0.41 Euro / liter.

5.5. Sensitivity analysis

The sensitivity analysis was carried out by calculating the percentage of each cost component to measure their contribution to the required selling price as can be seen in the table below.

<table>
<thead>
<tr>
<th>Plant capacity (ton/year)</th>
<th>125,000 US</th>
<th>125,000 US bioethanol</th>
<th>80,000 US</th>
<th>80,000 US bioethanol</th>
<th>8,000 US</th>
<th>8,000 US bioethanol</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Annual Variable Cost</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Raw Material</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Waste cooking oil</td>
<td>87.7</td>
<td>83.7</td>
<td>66.1</td>
<td>63.7</td>
<td>30.0</td>
<td>29.3</td>
</tr>
<tr>
<td>Methanol</td>
<td>14.2</td>
<td>10.8</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Bioethanol</td>
<td>16.0</td>
<td>12.3</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Start Up</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Methanol / bio-ethanol</td>
<td>4.8E-02</td>
<td>5.5E-02</td>
<td>3.6E-04</td>
<td>4.1E-02</td>
<td>1.6E-04</td>
<td>1.9E-04</td>
</tr>
<tr>
<td>Propane</td>
<td>1.5E-02</td>
<td>1.4E-02</td>
<td>1.1E-04</td>
<td>4.1E-02</td>
<td>4.7E-05</td>
<td>4.6E-05</td>
</tr>
<tr>
<td>Utilities</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Electricity</td>
<td>11.2</td>
<td>10.7</td>
<td>8.3</td>
<td>8.0</td>
<td>4.0</td>
<td>3.9</td>
</tr>
<tr>
<td>Cooling water</td>
<td>0.3</td>
<td>0.3</td>
<td>0.3</td>
<td>0.3</td>
<td>0.1</td>
<td>0.1</td>
</tr>
<tr>
<td><strong>By-product Credit</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Glycerol</td>
<td>53.6</td>
<td>51.2</td>
<td>24.5</td>
<td>23.6</td>
<td>17.9</td>
<td>17.4</td>
</tr>
<tr>
<td><strong>Fixed Cost</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Operating Labor</td>
<td>3.4</td>
<td>3.3</td>
<td>4.0</td>
<td>3.9</td>
<td>17.9</td>
<td>17.5</td>
</tr>
<tr>
<td>Maintenance</td>
<td>2.5</td>
<td>2.6</td>
<td>2.3</td>
<td>2.4</td>
<td>10.2</td>
<td>9.9</td>
</tr>
<tr>
<td>Plant Overhead</td>
<td>3.3</td>
<td>3.1</td>
<td>3.7</td>
<td>3.6</td>
<td>16.4</td>
<td>16.0</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
<td>1.1</td>
<td>1.1</td>
<td>1.0</td>
<td>1.0</td>
<td>1.1</td>
<td>1.1</td>
</tr>
<tr>
<td>Capital Charges</td>
<td>25.0</td>
<td>25.4</td>
<td>23.4</td>
<td>23.9</td>
<td>28.5</td>
<td>29.2</td>
</tr>
<tr>
<td>S, G &amp; A</td>
<td>4.8</td>
<td>4.8</td>
<td>4.8</td>
<td>4.8</td>
<td>4.8</td>
<td>4.8</td>
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<tr>
<td>% RSP</td>
<td>100</td>
<td>100</td>
<td>100</td>
<td>100</td>
<td>100</td>
<td>100</td>
</tr>
</tbody>
</table>

From the table above, it can be concluded that sensitive key factor for biodiesel plant using supercritical transesterification are: capital cost (charges), glycerol as by-product, and the price of raw material (waste cooking oil). The plant capacity changes / shifts the sensitive key factor dominantly from the price of raw material to operating labor, maintenance and plant overhead.

The smaller the plant capacity, the less influence of raw material, glycerol and capital cost contributes on the feasibility of the plant. Consequently, the smaller the capacity, the more sensitive the operating labor, maintenance and overhead cost have on the required selling price of biodiesel.
Table 17. Sensitive key factors for the biodiesel plants in the Netherlands

<table>
<thead>
<tr>
<th>Plant capacity (ton/year)</th>
<th>125,000</th>
<th>80,000</th>
<th>8,000</th>
</tr>
</thead>
<tbody>
<tr>
<td>Annual Variable Cost</td>
<td>% of RSP</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Raw Material</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Waste cooking oil</td>
<td>87.6</td>
<td>74.5</td>
<td>43.0</td>
</tr>
<tr>
<td>Methanol</td>
<td>7.7</td>
<td>6.6</td>
<td>3.8</td>
</tr>
<tr>
<td>Bioethanol</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Start Up</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Methanol / bio-ethanol</td>
<td>1.2E-01</td>
<td>2.2E-02</td>
<td>1.3E-04</td>
</tr>
<tr>
<td>Propane</td>
<td>3.5E-02</td>
<td>6.4E-03</td>
<td>3.7E-05</td>
</tr>
<tr>
<td>Utilities</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Electricity</td>
<td>6.0</td>
<td>5.0</td>
<td>3.1</td>
</tr>
<tr>
<td>Cooling water</td>
<td>0.3</td>
<td>0.2</td>
<td>0.1</td>
</tr>
<tr>
<td>By-product Credit</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Glycerol</td>
<td>29.0</td>
<td>15.0</td>
<td>13.8</td>
</tr>
<tr>
<td>Fixed Cost</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Operating Labor</td>
<td>1.9</td>
<td>2.4</td>
<td>13.9</td>
</tr>
<tr>
<td>Maintenance</td>
<td>1.4</td>
<td>1.4</td>
<td>7.9</td>
</tr>
<tr>
<td>Plant Overhead</td>
<td>1.8</td>
<td>2.2</td>
<td>12.7</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
<td>0.6</td>
<td>0.6</td>
<td>0.8</td>
</tr>
<tr>
<td>Capital Charges</td>
<td>17.1</td>
<td>17.3</td>
<td>23.8</td>
</tr>
<tr>
<td>S, G &amp; A</td>
<td>4.8</td>
<td>4.8</td>
<td>4.8</td>
</tr>
<tr>
<td>% RSP</td>
<td>100</td>
<td>100</td>
<td>100</td>
</tr>
</tbody>
</table>

From the table above, it can be concluded that sensitive key factor for biodiesel plant using supercritical transesterification in the Netherlands are: raw material cost and capital cost (charges). The biodiesel price is also influenced by the price of glycerol as by product. As seen before, the smaller the plant capacity, the more sensitive the fixed cost (operating labor, maintenance and overhead) contribution on the required selling price of biodiesel.
6. Results

The strategy behind the process design: minimizing the numbers of compressor, because of high capital cost and high operating cost. This is achieved by keeping all the flow ideally in liquid form. The lowest pressure is 5 bar to accommodate the propane gas in liquid phase at 40 °C. Consequently this is also the lowest temperature and the highest operating temperature is 280 °C. After process simulation, a compressor is inevitably needed to compress the gas phase of methanol from the vacuum flash evaporator. The high operating temperature of supercritical transesterification is achieved by using the hot recycle stream of methanol and propane in heat exchanger. This heat coupling is effective to decrease the operating cost.

The reaction as mentioned before was inserted in the Aspen software simulation as reaction input from the experimental results of Saka and Kusdiana (2001). A novel approach for the separation technique was carried out. This involved 2 flash tanks, one with 5 bar pressure and the other with vacuum condition. This approach was proven to be advantage for the cost reduction, since capital and operating cost of normal and vacuum distillation columns are considerable high. The process conditions for the separation step were chosen based on the result of sensitivity analysis for process design in the Aspen software.

High heat loss (6 MW) in the cooling tower B20 is inevitably occurred, this in agreement with the total amount of work applied to the system (6 MW). The heat loss can not be reduced, due to the limitation in the software unit models. The low pressure steam can not be effectively utilized to useful electric current. Based on the computer simulation, maximum power from low pressure steam turbine was 340 kW, which is a tiny amount, compared with total 6 MW energy requirements. In reality, recent low pressure steam turbines can generate more power than in Aspen turbine models.

It can be concluded that biodiesel by supercritical transesterification is technically feasible due to high yield, high purity biodiesel (99.8 %), high purity glycerol (97%) and high methanol and propane recovery (99.3%) attained in the process.

There are a lot of economic cost calculation methods in the world. One which is useful is the method of calculation for the required selling price. This method was described previously and given as an obligatory course in Eindhoven University of Technology by experienced former ABB Lummus R&D director Prof. Ir. F.M. Dautzenberg (2003). The Net Present value of the project was calculated based on the method of Coulson and Richardson (1998).

From the cost calculation for the 15 years project life (excluding 2 years construction time), it can be concluded that biodiesel by supercritical transesterification plants is economically feasible. The production of 125,000 ton per year biodiesel can be economically purchased at minimum price of US $ 0.20 / liter. For 80,000 ton/year the price is US $ 0.27 / liter. The required selling price is often also called break even price.
From figure above, it can be concluded that the biodiesel produced via supercritical transesterification which use no catalyst and no pre-treatment step is economically feasible since the biodiesel can be sold as low as 20 US cent per liter for the 125,000 ton/year capacity.

Biodiesel produced from small plant (8,000 ton/year) in the Netherlands is the most expensive compared with the others.

For all capacities, biodiesel price in the Netherlands is more expensive than in the United States. This can be attribute to the high price of waste cooking oil; 30 euro cent per liter (Rice et al, 1998) which is equal to 37 US cent/liter. The price of the waste cooking oil in the United States is 20 US cent / liter (Zhang et al, 2003).

Bio-ethanol hardly influences the biodiesel price as it only 1 US$ cent more expensive per liter. This can be attributed to considerably small sensitivity of the methanol or bio-ethanol as the reactants ranging between 5.6 - 16 % (table 16) although the bio-ethanol price is more expensive compared with methanol (bio-ethanol price US $ 355 /ton, methanol price US $ 300 /ton, appendix 8).
From the figure above, it can be seen that the biodiesel produced via non-catalytic way (55 US cent per liter) can compete with the industrial way produced biodiesel (Zhang 01 and 02 bars) (Zhang et al, 2003) as it is cheaper to sell. The most method used in biodiesel industries is alkali catalyst with acid pretreatment step prior to main reaction.

The high cost of the biodiesel (72 US cent per liter, second bar from the left) can be attributed to the high cost of the fresh vegetable oil. Whereas the required selling price of 74 US cent per liter is due to the additional pre-treatment step to reduce the free fatty acid of the waste cooking oil although the cost of the waste cooking oil is lower than fresh vegetable oil. The pre-treatment step increases the capital cost of the plant, consequently lead to higher price of biodiesel.

Zhang et al (2003) also carried out process design and economic feasibility with acid catalyst process which requires longer reaction time from waste cooking oil feed. It has advantage of insensitive to free fatty acids in the waste cooking oil. From the figure above, it is clear that biodiesel produced non-catalytically is 1 US cent more expensive than biodiesel produced with acid catalyst. Further comments can not be made because the literature does not show the cost calculation for net present value, and it is unclear the project life of the biodiesel plants.
From figure 12, it is clear that the investment of biodiesel plants is paid off in the 5th year. The highest the plant capacity, the highest the net cash flow attained from the production project. This can be attributed to higher sales attained in the big capacity production plants. It can be concluded that the plant capacity influences the net present value of the biodiesel plants as also can be seen in the table 18.

The higher cost of biodiesel in the Netherlands cause higher net cash flow that led to higher net present value. From figure above, it can be seen that the net present value of 80,000 ton/year plant capacity in the Netherlands is similar to the one of 125,000 ton/year in the United States. As mentioned before, the high cost of waste cooking oil led to high break even price of biodiesel in the Netherlands, which led to higher net present value due to higher net cash flow attained.
For big plants in the United States, the major cost contribution for the biodiesel price is the price of the raw material (64 - 88%) and capital charges (17-25%). Glycerol as the by-product also influences the price of the biodiesel as it contributes 29-54% reduction in the 125 kton/year capacity. Whereas in 80 kton/year capacity, the glycerol contributes 15-25% reduction of the biodiesel price.

For small plant capacity, the contribution of operating labor, maintenance, overhead become higher compared with bigger capacities.

This means that plant capacity changes the sensitive factors for cost contribution.
7. Conclusion

From the process design, economic feasibility and sensitivity analysis, it can be concluded that:
- biodiesel plant by supercritical transesterification is technically feasible,
- high purity biodiesel and glycerol as by-product attained
- biodiesel plant by supercritical transesterification is economically feasible,
- the sensitive / major contributors for the price: raw material price, plant capacity, glycerol price and capital cost
- the plant capacity changes the sensitive factors towards the human labor cost

8. Recommendation

- the high heat loss in the cooling tower is problematic although it did not influence the feasibility of the plant, in the future it should be tackled by modifying the existing turbine models in the Aspen software by i.e. re-programming the simulation unit with Mathlab for instance
- the exact supercritical bio-ethanol transesterification could not be simulated in the Aspen Plus software due to lack of components which can represent the ethyl esters. It is useful if Aspen can offer a great range of opportunity to represent this component
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**Internet sources**

a. www.energycrisis.com ; description of some experts forecast on energy crisis
b. http://en.wikipedia.org/wiki/Hubbert_peak#Peak_prediction; description of peak oil and natural gas, and their estimated depletion
c. www.scientificpsychic.com; properties of fatty acids and sources
### 1. The world consumption of oil and CO₂ emission

<table>
<thead>
<tr>
<th>HDI rank</th>
<th>Human Development Index (HDI) (2002)</th>
<th>population (million) (2002)</th>
<th>CO₂ emissions (ton/person)*</th>
<th>total CO₂ emissions (kiloton)</th>
<th>Oil consumption (kg/person)</th>
<th>Total oil consumption (kiloton)</th>
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</thead>
<tbody>
<tr>
<td><strong>High human development (rank 1-55)</strong></td>
<td></td>
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<tr>
<td>1</td>
<td>Norway</td>
<td>0.956</td>
<td>4.5</td>
<td>11.1</td>
<td>49950</td>
<td>6,019</td>
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<tr>
<td>2</td>
<td>Sweden</td>
<td>0.946</td>
<td>8.9</td>
<td>5.3</td>
<td>47170</td>
<td>5,693</td>
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<td>3</td>
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<td>8</td>
<td>United States</td>
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<td>9</td>
<td>Japan</td>
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<td>Germany</td>
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<td><strong>Medium human development (rank 56-141)</strong></td>
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<td>Russian Federation</td>
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<td>1.4</td>
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<td>Thailand</td>
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<td>1165</td>
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</table>
* based on year 2000 data (www.undp.org)  
** based on year 1999 data (www.worldbank.org)
### 2. Biodiesel standards

<table>
<thead>
<tr>
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<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
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</thead>
<tbody>
<tr>
<td>Density at 15°C</td>
<td>g/cm³</td>
<td>0.86 - 0.90</td>
<td>0.875 - 0.90</td>
<td>/</td>
<td>0.86 - 0.90</td>
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<tr>
<td>Viscosity at 40°C</td>
<td>mm²/s</td>
<td>6.5 - 9.0 (20°C)</td>
<td>3.5 - 5.0</td>
<td>1.9 - 6.0</td>
<td>3.50 - 5.00</td>
</tr>
<tr>
<td>Flash point</td>
<td>°C (°F)</td>
<td>min. 55 (131)</td>
<td>min. 110 (230)</td>
<td>min. 100 (212)</td>
<td>min. 120 (248)</td>
</tr>
<tr>
<td>CFPP</td>
<td>°C (°F)</td>
<td>max. 0 (32)</td>
<td>max. 0 (32)</td>
<td>/</td>
<td>/</td>
</tr>
<tr>
<td>Total sulphur</td>
<td>mg/kg</td>
<td>max. 200</td>
<td>max. 100</td>
<td>max. 500</td>
<td>max. 10.0</td>
</tr>
<tr>
<td>Conradson (CCR) at 100%</td>
<td>% mass</td>
<td>max. 0.1</td>
<td>max. 0.05</td>
<td>max. 0.05</td>
<td>/</td>
</tr>
<tr>
<td>at 10%</td>
<td>% mass</td>
<td>/</td>
<td>/</td>
<td>/</td>
<td>max. 0.30</td>
</tr>
<tr>
<td>Cetane number</td>
<td>-</td>
<td>min. 48</td>
<td>min. 49</td>
<td>min. 40</td>
<td>min. 51</td>
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<tr>
<td>Sulfated ash content</td>
<td>% mass</td>
<td>max. 0.02</td>
<td>max. 0.03</td>
<td>max. 0.02</td>
<td>max. 0.02</td>
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<tr>
<td>Water content</td>
<td>mg/kg</td>
<td>free of deposited water</td>
<td>max. 300</td>
<td>/</td>
<td>max. 500</td>
</tr>
<tr>
<td>Water &amp; sediment</td>
<td>vol. %</td>
<td>/</td>
<td>/</td>
<td>max. 0.05</td>
<td>/</td>
</tr>
<tr>
<td>Total contamination</td>
<td>mg/kg</td>
<td>/</td>
<td>max. 20</td>
<td>/</td>
<td>max. 24</td>
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<tr>
<td>Copper corrosion (3 hs, 50°C)</td>
<td>degree of Corrosion</td>
<td>/</td>
<td>1</td>
<td>No. 3b max.</td>
<td>1</td>
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<tr>
<td>Neutralisation value</td>
<td>mg</td>
<td>max. 1</td>
<td>max. 0.5</td>
<td>max. 0.8</td>
<td>max. 0.50</td>
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<tr>
<td>Oxidation stability</td>
<td>h</td>
<td>/</td>
<td>/</td>
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<td>min. 6.0</td>
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<td>Methanol content</td>
<td>% mass</td>
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<td>max. 0.3</td>
<td>max. 0.2</td>
<td>max. 0.20</td>
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<td>Ester content</td>
<td>% mass</td>
<td>/</td>
<td>/</td>
<td>/</td>
<td>min. 96.5</td>
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<tr>
<td>Monoglycerides</td>
<td>% mass</td>
<td>/</td>
<td>max. 0.8</td>
<td>/</td>
<td>max. 0.80</td>
</tr>
<tr>
<td>Diglycerides</td>
<td>% mass</td>
<td>/</td>
<td>max. 0.4</td>
<td>/</td>
<td>max. 0.20</td>
</tr>
<tr>
<td>Triglycerides</td>
<td>% mass</td>
<td>/</td>
<td>max. 0.4</td>
<td>/</td>
<td>max. 0.20</td>
</tr>
<tr>
<td>Free glycerine</td>
<td>% mass</td>
<td>max. 0.03</td>
<td>max. 0.02</td>
<td>max. 0.02</td>
<td>max. 0.02</td>
</tr>
<tr>
<td>Total glycerine</td>
<td>% mass</td>
<td>max. 0.25</td>
<td>max. 0.25</td>
<td>max. 0.24</td>
<td>max. 0.25</td>
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<tr>
<td>Iodine value</td>
<td>/</td>
<td>max. 115</td>
<td>/</td>
<td>/</td>
<td>max. 120</td>
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<tr>
<td>Linolenic acid ME</td>
<td>% mass</td>
<td>/</td>
<td>/</td>
<td>/</td>
<td>max. 12.0</td>
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<td>Polyunsaturated (&gt;=4db)</td>
<td>% mass</td>
<td>/</td>
<td>/</td>
<td>/</td>
<td>max. 1</td>
</tr>
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<td>Phosphorus content</td>
<td>mg/kg</td>
<td>/</td>
<td>max. 10</td>
<td>/</td>
<td>max. 10.0</td>
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<tr>
<td>Alkaline content (Na+K)</td>
<td>mg/kg</td>
<td>/</td>
<td>max. 5</td>
<td>/</td>
<td>max. 5.0</td>
</tr>
<tr>
<td>Alkaline earth metals (Ca + Mg)</td>
<td>mg/kg</td>
<td>/</td>
<td>/</td>
<td>/</td>
<td>max. 5.0</td>
</tr>
</tbody>
</table>

1) the world's first BioDiesel standard, ONORM C1190 (Feb 1991)

2) depending on the national appendix to EN 14214
3. Waste cooking oil and animal fat volume calculation
source: Dutch Biro of Statistics (www.cbs.nl)

<table>
<thead>
<tr>
<th>Year</th>
<th>Tax of collected waste cooking oil (million euro)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1999</td>
<td>78</td>
</tr>
<tr>
<td>2000</td>
<td>95</td>
</tr>
<tr>
<td>2001</td>
<td>116</td>
</tr>
<tr>
<td>average</td>
<td>96</td>
</tr>
</tbody>
</table>

per year

waste cooking oil volume = \( \frac{96 \times 10^6 \text{euro} \times 1 \text{(liter)dm}^3 \times 10^{-3} \text{m}^3 \times 953 \text{kg}}{0.20 \text{euro} \times 1 \text{dm}^3 \times 1 \text{m}^3} = 457440 \times 10^3 \text{kg/year} \)

Waste cooking oil volume in the Netherlands = 457,440 ton / year

4. Reaction kinetics calculation

The reaction kinetics was based on the supercritical experimental works of Kusdiana and Saka (2001) and used as the parameters for the transesterification reaction.

The transesterification reaction is essential for the biodiesel production:

Triglyceride (Tg) + 3 Methanol (MOH) ↔ 3 Methyl Esters (ME) + Glycerol (GI)

Tg refers to waste cooking oil. Three species can be defined as methyl esters (ME), glycerol (GI) and unmethyl esterified compounds (uME) which include triglyceride, diglyceride, monoglyceride and unreacted free fatty acids.

The reaction is assumed to proceed as a first order reaction, as a function of the decrease of molar flow of triglyceride and volume in a tubular plug flow reactor as the following

\[ r_{ME} = -\frac{dF_{Tg}}{dV} \]  \hspace{1cm} (1)

Equation 1 can be modified to

\[ r_{ME} = -\frac{dF_{uME}}{dV} \]  \hspace{1cm} (2)

For a first order reaction, the rate of reaction can be written as the product of reaction rate \( k \) (s\(^{-1}\)) and the concentration of unconverted Methyl esters \([uME]\) as following

\[ r_{ME} = -k[uME] \]  \hspace{1cm} (3)

Since the volumetric flow rate \( v_0 \) is constant, the equation can be rearranged as

\[ r_{ME} = \frac{dF_{uME}}{dV} = \frac{d([uME]v_0)}{dv} = v_0 \frac{d[uME]}{dV} \]  \hspace{1cm} (4)
Where $k$ is the reaction rate constant ($s^{-1}$) and can be rewritten to

$$r_{ME} = -k [uME] = v_0 \frac{d[uME]}{dV}$$

(5)

or

$$k = -\frac{v_0}{dV} \frac{d[uMe]}{[uME]}$$

(6)

$$k.dV = -v_0 \frac{d[uME]}{[uME]}$$

(7)

Assuming that at the entrance of the reactor when $V = 0$ and the initial concentration of $uMe$ at the entrance of reactor $uMe_0$ and it is reacted to $uMe$ at the reactor outlet (Fogler, 1986),

$$k.\int_{0}^{V} dV = -v_0 \int_{uMe_0}^{uMe} \frac{d[uME]}{[uME]}$$

(8)

the integration gives

$$k = \frac{v_0}{V} \ln \frac{uME_0}{uME}$$

(9)

and since\[ \frac{v_0 (m^3/s)}{V (m^3)} = \frac{1}{\theta} (s^{-1}), \]

$$k = \frac{\ln [uME_0] - \ln [uME]}{\theta}$$

(10)

The reactor was modeled as a plug flow reactor (RPlug) in Aspen Plus with the properties as in the table 2. The transesterification reaction, activation energy and other kinetic parameters were entered as the reaction input. The activation energy was obtained by the calculation and estimation of the data of the Arrhenius plot of Kusdiana and Saka (2001).

**Arrhenius plot translation to reaction kinetics**

Based on the experimental results of Kusdiana and Saka (2001),

<table>
<thead>
<tr>
<th>Temperature</th>
<th>$^\circ$C</th>
<th>$^\circ$K</th>
<th>$1/T$</th>
<th>$k$ (s$^{-1}$)</th>
<th>$\ln k$</th>
<th>$P$ Mpa</th>
<th>$P$ bar</th>
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</thead>
<tbody>
<tr>
<td>200</td>
<td>230</td>
<td>270</td>
<td>211 x10$^3$</td>
<td>$2 \times 10^{-4}$</td>
<td>-8.52</td>
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<td>70</td>
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<tr>
<td></td>
<td>230</td>
<td>270</td>
<td>199 x10$^3$</td>
<td>$3 \times 10^{-4}$</td>
<td>-8.11</td>
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<td>90</td>
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<tr>
<td></td>
<td>270</td>
<td>270</td>
<td>184 x10$^3$</td>
<td>$7 \times 10^{-5}$</td>
<td>-7.26</td>
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<td>120</td>
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</table>
And the Arhenius plot can be seen below

From this plot, the activation energy of the transesterification can be calculated from the linear regression, thus:

\[ y = -4628.3x + 1.2055 \]
\[ R^2 = 0.9753 \]

<table>
<thead>
<tr>
<th>x (1/K)</th>
<th>y</th>
<th>ln K</th>
<th>slope</th>
<th>Energy of Activation</th>
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</thead>
<tbody>
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<td>0.003289</td>
<td>-14.02</td>
<td>-4628.3</td>
<td>38481.5 J / mol</td>
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<tr>
<td>0.001808</td>
<td>-7.164</td>
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<tr>
<td>2.1E-03</td>
<td>-8.579</td>
<td>-4628.3</td>
<td>38481.5 kJ / kmol</td>
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<tr>
<td>1.84E-03</td>
<td>-7.318</td>
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</table>

Reaction input in Aspen Plus model:
On the process window, go to reactions, and then setup the transesterification as the first order reaction. Choose [C] basis in molarity, then the kinetic expression is:

\[ k = A_o \cdot e^{-\frac{E}{RT}} \]

\( A_o \) = pre-exponential factor in SI unit
\( T \) = operating temperature
\( E \) = activation energy (kJ/kmol)

From the reaction rate constant, activation energy and operating temperature of 280 °C, the \( A_o \) was calculated: 1.159 s\(^{-1}\).
These numbers are inserted in the Aspen Plus reaction parameters.
5. Thermodynamic properties estimation of oleic acid and methyl esters

The oleic acid heat of formation

source
book: The properties of gases and liquids, by Reid, Prausnitz and Sherwood page 257, 261-265
Calculation for the heat of formation according to Franklin Method (The properties of gases and liquids, by Reid, Prausnitz and Sherwood)

<table>
<thead>
<tr>
<th>group</th>
<th>amount</th>
<th>Delta H (kcal/gmol)</th>
<th>total delta H</th>
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</thead>
<tbody>
<tr>
<td>CH3</td>
<td>3</td>
<td>-10.12</td>
<td>-30.36</td>
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<tr>
<td>CH2</td>
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<tr>
<td>cis HC=CH</td>
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<td>18.88</td>
<td>56.64</td>
</tr>
<tr>
<td>CH</td>
<td>1</td>
<td>-1.09</td>
<td>-1.09</td>
</tr>
<tr>
<td>COO</td>
<td>3</td>
<td>-79.8</td>
<td>-239.4</td>
</tr>
</tbody>
</table>

\[ \text{Delta H} = -431.13 \text{ kcal/gmol} \]

The oleic acid gibbs free energy

source
book: The properties of gases and liquids, by Reid, Prausnitz and Sherwood page 278-286
van Krevelen and Chermin method
Equation 7-6.1 \[ \Delta G = A + B \times T \]
T reaction = 280 oC = 553 oK

<table>
<thead>
<tr>
<th>chemical group</th>
<th>amount</th>
<th>A</th>
<th>B</th>
<th>sum A</th>
<th>sum B</th>
</tr>
</thead>
<tbody>
<tr>
<td>CH3</td>
<td>3</td>
<td>-10.943</td>
<td>2.215</td>
<td>-32.829</td>
<td>0.066</td>
</tr>
<tr>
<td>CH2</td>
<td>44</td>
<td>-5.193</td>
<td>2.430</td>
<td>-228.492</td>
<td>1.069</td>
</tr>
<tr>
<td>cis HC=CH</td>
<td>3</td>
<td>17.663</td>
<td>1.965</td>
<td>52.989</td>
<td>0.059</td>
</tr>
<tr>
<td>CH</td>
<td>1</td>
<td>-0.705</td>
<td>2.910</td>
<td>-0.705</td>
<td>0.029</td>
</tr>
<tr>
<td>COO</td>
<td>3</td>
<td>-92.620</td>
<td>2.610</td>
<td>-277.860</td>
<td>0.078</td>
</tr>
</tbody>
</table>

\[ \Delta G = 233.11 \text{ kcal/gmol} \]

Delta G

Methyl esters (biodiesel) heat of formation and gibbs free energy

book: The properties of gases and liquids, by Reid et al page 257, 261-265
Calculation for the heat of formation according to Franklin Method (The properties of gases and liquids, by Reid, Prausnitz and Sherwood)

<table>
<thead>
<tr>
<th>group</th>
<th>amount</th>
<th>Delta H (kcal/gmol)</th>
<th>total delta H</th>
</tr>
</thead>
<tbody>
<tr>
<td>CH3</td>
<td>2</td>
<td>-10.12</td>
<td>-20.24</td>
</tr>
<tr>
<td>CH2</td>
<td>14</td>
<td>-4.93</td>
<td>-69.02</td>
</tr>
<tr>
<td>cis HC=CH</td>
<td>1</td>
<td>18.88</td>
<td>18.88</td>
</tr>
<tr>
<td>CH</td>
<td>0</td>
<td>-1.09</td>
<td>0</td>
</tr>
<tr>
<td>COO</td>
<td>1</td>
<td>-79.8</td>
<td>-79.8</td>
</tr>
</tbody>
</table>

\[ \text{Delta H} = -150.18 \text{ kcal/gmol} \]
Calculation for the **Gibbs energy of formation**

book: The properties of gases and liquids, by Reid, Prausnitz and Sherwood

page 278-286

van Krevelen and Chermin method

<table>
<thead>
<tr>
<th>Equation 7-6.1</th>
<th>Delta G = A + B*T</th>
<th>T reaction = 280 °C = 553 oK</th>
</tr>
</thead>
<tbody>
<tr>
<td>chemical group</td>
<td>amount</td>
<td>A</td>
</tr>
<tr>
<td>CH3</td>
<td>2</td>
<td>-10.943</td>
</tr>
<tr>
<td>CH2</td>
<td>14</td>
<td>-5.193</td>
</tr>
<tr>
<td>cis HC=CH</td>
<td>1</td>
<td>17.663</td>
</tr>
<tr>
<td>CH</td>
<td>0</td>
<td>-0.705</td>
</tr>
<tr>
<td>COO</td>
<td>1</td>
<td>-92.620</td>
</tr>
</tbody>
</table>

**Delta G** 68.38 kcal/gmol

6. Reactor design calculation

Dynamic viscosity was estimated by the method of van Velzen *et al* as can be seen below.

Perry's Chemical Engineer's Handbook Page 2-365

Liquid viscosity estimation of Van Velzen et al

<table>
<thead>
<tr>
<th>component</th>
<th>N*</th>
<th>To (K)</th>
<th>log Niu</th>
<th>Tcritical (K)</th>
<th>Pcritical (bar)</th>
<th>viscosity (Pa.s)</th>
<th>Viscosity (kg /ms)</th>
</tr>
</thead>
<tbody>
<tr>
<td>oleic acid</td>
<td>5.75</td>
<td>203.2</td>
<td>-2.98</td>
<td></td>
<td></td>
<td>0.0010</td>
<td>1.0E-03</td>
</tr>
<tr>
<td>methanol</td>
<td>11.33</td>
<td>309.3</td>
<td>-3.29</td>
<td>512.1</td>
<td>80.9</td>
<td>0.0005</td>
<td>5.1E-04</td>
</tr>
<tr>
<td>propane</td>
<td>3.68</td>
<td>149.2</td>
<td>-3.08</td>
<td>369.9</td>
<td>42.5</td>
<td>0.0008</td>
<td>8.4E-04</td>
</tr>
</tbody>
</table>

total dynamic viscosity 0.0005 Kg / m s

Residence time 9 minutes

Re = 10000 for turbulent flow

Checking the Reynolds number of adiabatic reactor

with ±10% accuracy (Sinnot, Coulson, 1998 page 277-281)

<table>
<thead>
<tr>
<th>Section of tubular reactor</th>
<th>beginning</th>
<th>end</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>280 °C</td>
<td>240 °C</td>
</tr>
<tr>
<td>Density (kg/m³)</td>
<td>315</td>
<td>579</td>
</tr>
<tr>
<td>Dynamic viscosity (kg/m.s)</td>
<td>5 x 10⁻⁴</td>
<td>6.75 x 10⁻⁴</td>
</tr>
<tr>
<td>v.D relation (m²/s)</td>
<td>1.73 x 10⁻²</td>
<td>1.73 x 10⁻²</td>
</tr>
<tr>
<td>Reynolds</td>
<td>10,000</td>
<td>14,840</td>
</tr>
<tr>
<td></td>
<td></td>
<td>12,142</td>
</tr>
</tbody>
</table>
For 125,000 ton / year capacity, the relation between reactor parameters can be seen below

<table>
<thead>
<tr>
<th>Internal diameter (m)</th>
<th>Thickness of tube (mm)</th>
<th>V (m/s)</th>
<th>Number of tubes</th>
<th>Nt</th>
<th>Lt (m)</th>
<th>Total volume of tubes (m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.010</td>
<td>0.70</td>
<td>1.731</td>
<td>202.1</td>
<td>202</td>
<td>953.56</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.020</td>
<td>1.39</td>
<td>0.865</td>
<td>101.0</td>
<td>101</td>
<td>476.78</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.030</td>
<td>2.09</td>
<td>0.571</td>
<td>67.4</td>
<td>68</td>
<td>314.74</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.040</td>
<td>2.78</td>
<td>0.428</td>
<td>50.5</td>
<td>51</td>
<td>236.05</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.050</td>
<td>3.48</td>
<td>0.341</td>
<td>40.4</td>
<td>41</td>
<td>187.92</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.060</td>
<td>4.17</td>
<td>0.286</td>
<td>33.7</td>
<td>34</td>
<td>157.37</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.070</td>
<td>4.87</td>
<td>0.246</td>
<td>28.9</td>
<td>29</td>
<td>135.55</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.080</td>
<td>5.57</td>
<td>0.210</td>
<td>25.3</td>
<td>26</td>
<td>115.76</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.090</td>
<td>6.26</td>
<td>0.188</td>
<td>22.5</td>
<td>23</td>
<td>103.39</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.100</td>
<td>6.96</td>
<td>0.166</td>
<td>20.2</td>
<td>21</td>
<td>91.72</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.110</td>
<td>7.65</td>
<td>0.152</td>
<td>18.4</td>
<td>19</td>
<td>83.78</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.120</td>
<td>8.35</td>
<td>0.143</td>
<td>16.8</td>
<td>17</td>
<td>78.68</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.130</td>
<td>9.04</td>
<td>0.129</td>
<td>15.5</td>
<td>16</td>
<td>71.23</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.140</td>
<td>9.74</td>
<td>0.119</td>
<td>14.4</td>
<td>15</td>
<td>65.52</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.150</td>
<td>10.43</td>
<td>0.111</td>
<td>13.5</td>
<td>14</td>
<td>61.15</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.160</td>
<td>11.13</td>
<td>0.105</td>
<td>12.6</td>
<td>13</td>
<td>57.88</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.170</td>
<td>11.83</td>
<td>0.101</td>
<td>11.9</td>
<td>12</td>
<td>55.54</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.180</td>
<td>12.52</td>
<td>0.098</td>
<td>11.2</td>
<td>11</td>
<td>54.05</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.190</td>
<td>13.22</td>
<td>0.088</td>
<td>10.6</td>
<td>11</td>
<td>48.51</td>
<td>4.207456</td>
</tr>
<tr>
<td>0.200</td>
<td>13.91</td>
<td>0.087</td>
<td>10.1</td>
<td>10</td>
<td>48.15</td>
<td>4.207456</td>
</tr>
</tbody>
</table>

Reactor construction (125,000 ton/year capacity):
21 parallel series
1 series contains of 8 tubes (internal diameter of 10cm, 7 mm thick) of 12 m length in series with 7 bends, then 21 of these series stack together to construct the reactor.

For 80,000 ton / year capacity, the relation between reactor parameters can be seen below

<table>
<thead>
<tr>
<th>Internal diameter (m)</th>
<th>Thickness of tube (mm)</th>
<th>V (m/s)</th>
<th>Number of tubes</th>
<th>Nt</th>
<th>Lt (m)</th>
<th>Total volume of tubes (m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.010</td>
<td>0.70</td>
<td>1.088</td>
<td>202.1</td>
<td>202</td>
<td>599.34</td>
<td>2.644528</td>
</tr>
<tr>
<td>0.020</td>
<td>1.39</td>
<td>0.544</td>
<td>101.0</td>
<td>101</td>
<td>299.67</td>
<td>2.644528</td>
</tr>
<tr>
<td>0.030</td>
<td>2.09</td>
<td>0.359</td>
<td>67.4</td>
<td>68</td>
<td>197.82</td>
<td>2.644528</td>
</tr>
<tr>
<td>0.040</td>
<td>2.78</td>
<td>0.269</td>
<td>50.5</td>
<td>51</td>
<td>148.37</td>
<td>2.644528</td>
</tr>
<tr>
<td>0.050</td>
<td>3.48</td>
<td>0.214</td>
<td>40.4</td>
<td>41</td>
<td>118.11</td>
<td>2.644528</td>
</tr>
<tr>
<td>0.060</td>
<td>4.17</td>
<td>0.180</td>
<td>33.7</td>
<td>34</td>
<td>98.91</td>
<td>2.644528</td>
</tr>
<tr>
<td>0.070</td>
<td>4.87</td>
<td>0.155</td>
<td>28.9</td>
<td>29</td>
<td>85.20</td>
<td>2.644528</td>
</tr>
<tr>
<td>0.080</td>
<td>5.57</td>
<td>0.132</td>
<td>25.3</td>
<td>26</td>
<td>72.76</td>
<td>2.644528</td>
</tr>
<tr>
<td>0.090</td>
<td>6.26</td>
<td>0.118</td>
<td>22.5</td>
<td>23</td>
<td>64.99</td>
<td>2.644528</td>
</tr>
<tr>
<td>0.100</td>
<td>6.96</td>
<td>0.105</td>
<td>20.2</td>
<td>21</td>
<td>57.65</td>
<td>2.644528</td>
</tr>
</tbody>
</table>
Reactor construction (80,000 ton/year capacity):
26 parallel series
1 series contains of 6 tubes (internal diameter of 8 cm, 6 mm thick) of 12 m length in series with 5 bends, then 26 of these series stack together to construct the reactor.

For 8,000 ton/year capacity, the relation between reactor parameters can be seen below

<table>
<thead>
<tr>
<th>Internal diameter (cm)</th>
<th>Thickness of tube (mm)</th>
<th>V (m/s)</th>
<th>Number of tubes</th>
<th>Nt</th>
<th>Lt (m)</th>
<th>Total volume of tubes (m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.7</td>
<td>1.721</td>
<td>12.9</td>
<td>13</td>
<td></td>
<td>948.43</td>
<td>0.269321</td>
</tr>
<tr>
<td>1.0</td>
<td>1.105</td>
<td>8.6</td>
<td>9</td>
<td></td>
<td>608.87</td>
<td>0.269321</td>
</tr>
<tr>
<td>1.3</td>
<td>0.799</td>
<td>6.5</td>
<td>7</td>
<td></td>
<td>440.34</td>
<td>0.269321</td>
</tr>
<tr>
<td>1.5</td>
<td>0.716</td>
<td>5.2</td>
<td>5</td>
<td></td>
<td>394.55</td>
<td>0.269321</td>
</tr>
<tr>
<td>2.0</td>
<td>0.622</td>
<td>4.3</td>
<td>4</td>
<td></td>
<td>342.49</td>
<td>0.269321</td>
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<td>0.457</td>
<td>3.7</td>
<td>4</td>
<td></td>
<td>251.62</td>
<td>0.269321</td>
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<tr>
<td>3.0</td>
<td>0.466</td>
<td>3.2</td>
<td>3</td>
<td></td>
<td>256.87</td>
<td>0.269321</td>
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<td>0.368</td>
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<td></td>
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<td>0.269321</td>
</tr>
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<td>3</td>
<td></td>
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<td></td>
<td>135.86</td>
<td>0.269321</td>
</tr>
<tr>
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<td>2.2</td>
<td>2</td>
<td></td>
<td>171.24</td>
<td>0.269321</td>
</tr>
<tr>
<td>5.5</td>
<td>0.265</td>
<td>2.0</td>
<td>2</td>
<td></td>
<td>145.91</td>
<td>0.269321</td>
</tr>
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<td>1.8</td>
<td>2</td>
<td></td>
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<td>0.269321</td>
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<td>2</td>
<td></td>
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<td>0.269321</td>
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<td>2</td>
<td></td>
<td>96.32</td>
<td>0.269321</td>
</tr>
<tr>
<td>7.5</td>
<td>0.155</td>
<td>1.5</td>
<td>2</td>
<td></td>
<td>85.33</td>
<td>0.269321</td>
</tr>
<tr>
<td>8.0</td>
<td>0.138</td>
<td>1.4</td>
<td>2</td>
<td></td>
<td>76.11</td>
<td>0.269321</td>
</tr>
<tr>
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<td>0.124</td>
<td>1.4</td>
<td>2</td>
<td></td>
<td>68.31</td>
<td>0.269321</td>
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<td>9.0</td>
<td>0.112</td>
<td>1.3</td>
<td>2</td>
<td></td>
<td>61.65</td>
<td>0.269321</td>
</tr>
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<td>0.056</td>
<td>1.3</td>
<td>4</td>
<td></td>
<td>30.82</td>
<td>0.269321</td>
</tr>
</tbody>
</table>

For the smallest capacity (8,000 ton/year) assumed that the maximum length is 40 meter and maximum number of tubes are 4, so 4 tubes with length 32 m which has internal diameter of 10 cm and 7 mm thick is good option to perform the reaction.

Reactor construction (8,000 kton/year capacity):
4 parallel series
1 series contains of 8 tubes (internal diameter of 10cm, 7 mm thick) of 4 m length in series with 7 bends, then 4 of these series stack together to construct the reactor.
7. Sensitivity Analysis Results

process conditions of the vacuum flash evaporator chosen from sensitivity analysis at 43 °C

<table>
<thead>
<tr>
<th>Pressure atm</th>
<th>Methanol in end product %</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.01</td>
<td>0.07</td>
</tr>
<tr>
<td>0.02</td>
<td>0.15</td>
</tr>
<tr>
<td>0.03</td>
<td>0.20</td>
</tr>
<tr>
<td>0.04</td>
<td>0.19</td>
</tr>
<tr>
<td>0.05</td>
<td>0.18</td>
</tr>
</tbody>
</table>

at pressure of 0.02 atm

<table>
<thead>
<tr>
<th>T in Celsius</th>
<th>Methanol in the end stream %</th>
</tr>
</thead>
<tbody>
<tr>
<td>35</td>
<td>0.21</td>
</tr>
<tr>
<td>36</td>
<td>0.21</td>
</tr>
<tr>
<td>36</td>
<td>0.20</td>
</tr>
<tr>
<td>37</td>
<td>0.20</td>
</tr>
<tr>
<td>38</td>
<td>0.20</td>
</tr>
<tr>
<td>38</td>
<td>0.19</td>
</tr>
<tr>
<td>39</td>
<td>0.19</td>
</tr>
<tr>
<td>40</td>
<td>0.19</td>
</tr>
<tr>
<td>40</td>
<td>0.18</td>
</tr>
<tr>
<td>41</td>
<td>0.18</td>
</tr>
<tr>
<td>42</td>
<td>0.17</td>
</tr>
<tr>
<td>42</td>
<td>0.17</td>
</tr>
<tr>
<td>43</td>
<td>0.17</td>
</tr>
<tr>
<td>44</td>
<td>0.17</td>
</tr>
<tr>
<td>44</td>
<td>0.16</td>
</tr>
<tr>
<td>45</td>
<td>0.16</td>
</tr>
</tbody>
</table>
8. Components prices

<table>
<thead>
<tr>
<th>Name</th>
<th>Unit</th>
<th>Price</th>
<th>Source</th>
</tr>
</thead>
<tbody>
<tr>
<td>Waste cooking oil</td>
<td>US$ / liter</td>
<td>0.20</td>
<td>0.18 US$ / liter waste cooking oil (Zhang, 2003) + 0.02 US$ / liter transportation cost.</td>
</tr>
<tr>
<td>Bio-ethanol</td>
<td>US$ / ton</td>
<td>355</td>
<td>16 pence / liter (brazil imported bioethanol) United Kingdom Department of Transport <a href="http://www.dft.gov.uk/stellent/groups/dft_roads/documents/page/dft_roads_024054-10.hcsp">http://www.dft.gov.uk/stellent/groups/dft_roads/documents/page/dft_roads_024054-10.hcsp</a> or 0.28 US$ / liter bio-ethanol density 0.789 kg / liter United Kingdom Department of Transport <a href="http://www.dft.gov.uk/stellent/groups/dft_roads/documents/page/dft_roads_024054-06.hcsp">www.dft.gov.uk/stellent/groups/dft_roads/documents/page/dft_roads_024054-06.hcsp</a></td>
</tr>
<tr>
<td>Methanol</td>
<td>US$ / ton</td>
<td>300</td>
<td><a href="http://www.methanex.com">www.methanex.com</a> valid through 31st July 2005</td>
</tr>
<tr>
<td>Propane</td>
<td>US$ / ton</td>
<td>1102</td>
<td><a href="http://www.uniongas.com">www.uniongas.com</a></td>
</tr>
<tr>
<td>Electricity (NL)</td>
<td>US$ / kWh</td>
<td>0.079</td>
<td>3.75 pence / kWh for extra large industrial prices (2003), source: report to DTI (Department of Transport and Industry UK) from Ilex energy consulting, Impact of the EU ETS on european electricity prices, July 2004 0.0375 GBP = 0.0675070 USD * cepci index (468.3 / 402) US$ 0.71 / kWh from January 2004 data of Department of Trade &amp; Industry United Kingdom <a href="http://www.dti.gov.uk/energy/inform/energy_prices/section5_mar04.shtml">http://www.dti.gov.uk/energy/inform/energy_prices/section5_mar04.shtml</a></td>
</tr>
<tr>
<td>Cooling water</td>
<td>US$ / ton</td>
<td>0.86</td>
<td>Increased by 4 % per year worldwide (result of NUS consulting group study) from average US water price US$ 0.764/m³, February 2002 Arizona Water Resource (<a href="http://ag.arizona.edu/AZWATER/awr/JanFeb02/news.html">http://ag.arizona.edu/AZWATER/awr/JanFeb02/news.html</a>) Remark: 0.0017 (Aspen IPE 11.1 default value)</td>
</tr>
<tr>
<td>Glycerol</td>
<td>US$ / ton</td>
<td>600</td>
<td><a href="http://www.chemsource.com">www.chemsource.com</a> 110 US$ / 3.78541 liter (1gallon US) glycerol <a href="http://www.chemdat.de">www.chemdat.de</a> (Merck chemical database) glycerol (for synthesis) density = 1.26 g / cm³</td>
</tr>
<tr>
<td>Year</td>
<td>CEPCI</td>
<td></td>
<td></td>
</tr>
<tr>
<td>----------</td>
<td>--------</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1998</td>
<td>389.5</td>
<td></td>
<td></td>
</tr>
<tr>
<td>2003</td>
<td>402.0</td>
<td></td>
<td></td>
</tr>
<tr>
<td>2004</td>
<td>444.2</td>
<td></td>
<td></td>
</tr>
<tr>
<td>March 2005</td>
<td>468.3</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

(Source: Chemical Engineering, Economic Indicators, July 2005)

Key process equipment typical sizes and prices
(Source: Dautzenberg, 2003 - ABB Lummus global, Eindhoven University of Technology)

**Compressors**
Basis: Centrifugal type without drivers

<table>
<thead>
<tr>
<th>Nominal Size, bhp</th>
<th>Purchase price, USGC 1998</th>
</tr>
</thead>
<tbody>
<tr>
<td>300</td>
<td>$300,000</td>
</tr>
<tr>
<td>3,000</td>
<td>$1,225,000</td>
</tr>
<tr>
<td>30,000</td>
<td>$50,000,000</td>
</tr>
</tbody>
</table>

Assumed that bhp = hp (electric), 1 kW = 0.746 hp (electric) (Perry's Handbook, 1998);
The total net power requirement is 2948 kW (125 kton/year capacity);
Hence the power required is 2948 x 0.746 hp = 2199 hp;
The price of the compressor = 1,225,000 $ x 468.3 / 389.5 = 1,472,831 US $.
9. Cooling tower design calculation
Source: Perry's Chemical Engineer's Handbook page 12-21
7th edition, 1997

**Natural-draft towers**

\[
D_t = \frac{(A \cdot Z_t)}{C_i \cdot C_t}
\]

- \(D_t\) = duty coefficient
- \(A\) = base area of tower (ft\(^2\))
- \(Z_t\) = height of tower (ft)
- \(C_i\) = performance coefficient or efficiency factor

\[
W' = \frac{90.59 \cdot \Delta h}{\Delta T} \cdot \sqrt{\Delta t + 0.3124 \cdot \Delta h}
\]

- \(W'\) = water load to tower (lb/h)
- \(\Delta h\) = change of total heat of air passing through tower
- \(\Delta T\) = change of temperature of water passing through tower
- \(\Delta t\) = difference between air temperature leaving the packing and inlet dry-bulb temperature

\[
T_i = 120 \, ^\circ C = 248 \, ^\circ F
\]
\[
T_o = 22 \, ^\circ C = 71.6 \, ^\circ F
\]
\[
\Delta T = 176.4 \, ^\circ F
\]

\[
t_1 = \frac{248 + 71.6}{2} = 159.8 \, ^\circ F \rightarrow h_1 = 40 \, \text{Btu/lb (from figure 12-3)}
\]
\[
t_2 = 57 \, ^\circ F \rightarrow h_2 = 21.3 \, \text{Btu/lb}
\]
\[
\Delta t = 102.8 \, ^\circ F, \Delta h = 18.7 \, \text{Btu/lb}
\]
\[
W' = \frac{90.59 \cdot 18.7}{176.4} \cdot \sqrt{102.8 + 0.3124 \cdot 18.7} = 100.1
\]
\[
D_t = \frac{39065.9 \, \text{lb/h} \text{ (converted from 17720 kg/h) cooling water}}{100.1} = 390.3
\]

\(C_i\) assumed as 5.0 and ratio of height to base diameter of 3:2 is normal,

\[
390.3 = \frac{\left(\frac{1}{4} \pi d^2 \cdot \sqrt{1.5 \, d}\right)}{(C_i \cdot \sqrt{C_t})} \rightarrow 872.7 = 0.25 \pi d^2 \cdot \sqrt{1.5 \, d} \rightarrow d = 16.5 \, \text{ft} = 5.1 \, \text{m}
\]

thus the base diameter of the tower = 5.1 m, and height of tower = 7.6 m
10. Purchase cost calculation for process equipments

Reactor (B1):
Construction: 21 parallel series
1 series contains of 8 tubes (internal diameter of 10 cm, 7 mm thick) of 12 m length in series with 7 bends, then 21 of these series stack together to construct the reactor.
For cost calculation:
100 m of 4 inch diameter cost 1,210 euro, so 1,210 x 12/100 x 21 x 8 x (468.3/402) = 28,416.7 euro.
The bends = 7 * 7 euro * 21 * (468.3/402) = 1,198.7 Euro
Isolation for the reactor:
From above, the series which contains 8 tubes in series = 8 x (10 cm+ 0.7cm+0.7 cm) = 91.2 cm = 0.912 m wide and 12 m length = 0.9 x 12 m = 10.8 m$^2$ x 83 euro (isolatiedikte 100 voor vlakke wanden) = 896.4 euro x 2 = 1792.8 euro
From the side view, 12 m x 0.114 m = 1.368 x 83 euro = 113.544 x 2 = 227.088 euro
From the other side view, 0.912 m x 0.114 m = 0.104 x 83 euro = 8.63 x 2 = 17.26 euro
Isolation = (1792.8 + 227.09 + 17.26) x (468.3/402) = 2373.13 euro x 21 = 49,835.73 euro
Total reactor purchase cost = 79,451.1 euro x 1.232 = 97,884 US $.

Flash evaporator (B16) with 1.13 m diameter and 2.39 m height; from DACE costbook a steel column with diameter of 1.5 m and height of 5 m costs 34,000 Euro = times the CEPCI index and time the conversion euro to dollar gives 78,074 US $.

Vacuum flash evaporator (B8) with 4.9 m diameter and 5.4 m height; steel tank of 10m height and 4 m diameter costs 85,000. Assumed this is similar, the purchase cost is 243,982 US $.

Settler tank (B11) with 2 m diameter and 4 m height; steel tank with the same diameter and 5 m height costs 37,000 Euro, then equal to 53,102 US$.

Heat exchanger 1 (B6) requires area of 1059 m$^2$, a heat exchanger constructed of stainless steel material (AISI 304) tubes (with each diameter of 2cm) with area of 1000 m$^2$ costs 262,000 Euro. With area conversion (price*1059/1000), index and conversion rate gives 398,204 US $.

Heat exchanger 2 (B15) requires area of 109 m$^2$, a heat exchanger constructed of stainless steel material (AISI 304) tubes (with each diameter of 2cm) with area of 100 m$^2$ costs 38,000 Euro. With area conversion, index and conversion rate gives 59,445 US $.

Heat exchanger 3 (B10) requires area of 100 m$^2$, a heat exchanger constructed of stainless steel material (AISI 304) tubes (with each diameter of 2cm) with area of 100 m$^2$ costs 38,000 Euro. With area conversion, index and conversion rate gives 54,537 US $.

Heat exchanger 4 (B13) requires area of 31 m$^2$, a heat exchanger constructed of stainless steel material (AISI 304) tubes (with each diameter of 2cm) with area of 30 m$^2$ costs 24,000 Euro. With area conversion, index and conversion rate gives 34,445 US $.

Cooling tower (B20) with 5.1 m diameter and 7.6 m height = 6 x 8 m times 1,000 Euro (one layer concrete industrial building / bouwkunde in dutch) times the index and conversion rate gives 118,272 US $.
Compressor (B14), assumed that bhp = hp (electric), 1 kW = 0.746 hp (electric) (Perry’s Handbook, 1998); The total net power requirement is 2948 kW (125 kton/year capacity); Hence the power required is 2948 x 0.746 hp = 2199 hp (also in appendix 8). The price of the compressor = 1,225,000 $ x 468.3 / 389.5 = 1,472,831 US $.

Pump 1 (B5) with required work duty of 284 kW; a pump of 140 kW costs 24,900 Euro, then 24,900*284/140, times the index and conversion rate to dollar = 58,842 US $.

Pump 2 (B2) with required work duty of 10 kW; a pump of 11 kW with the highest pump capacity, stainless steel internal parts (AISI 316) (Dace costbook, 2003) costs 9,100 Euro, then times the index and conversion rate to dollar = 13,060 US $.

Pump 3 (B3) with required work duty of 1 kW; a pump of 1.1 kW with the highest pump capacity, stainless steel internal parts (AISI 316) costs 5,200 Euro, then times the index and conversion rate to dollar = 6,785 US $.

Pump 4 (B12) with required work duty of 1 kW; a pump of 1.1 kW with the highest pump capacity, stainless steel internal parts (AISI 316) costs 5,200 Euro, then times the index and conversion rate to dollar = 6,785 US $.

Pump 5 (B19) with required work duty of 1 kW; a pump of 1.1 kW with the highest pump capacity, stainless steel internal parts (AISI 316) costs 5,200 Euro, then times the index and conversion rate to dollar = 6,785 US $.

The calculation method for smaller capacities

For rapid cost calculation, the usual method is to use the six-tenth rule of thumb for the relation between the capital cost and plant capacity as can be seen below (Sinnot, Coulson, 1998).

\[ C_2 = C_1 \left( \frac{S_2}{S_1} \right)^{n-0.6} \]

where

\[ C_2 = \text{capital cost of project with capacity } S_2 \]
\[ C_1 = \text{capital cost of project with capacity } S_1 \]

Dividing the capital cost with the Lang factor (5), the capital cost of six-tenth rule is assumed to be the same as the purchase equipment cost.

The purchase cost of equipment can be also calculated directly from the costbook (with 30% accuracy depending on the supplier as stated in the DACE costbook, 2003).
The purchase equipment cost calculated by six-tenth rule and costbook can be seen below.

<table>
<thead>
<tr>
<th>Unit</th>
<th>Method</th>
<th>costbook</th>
<th>six-tenth rule</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reactor:</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>steel</td>
<td></td>
<td>32,960</td>
<td></td>
</tr>
<tr>
<td>isolation</td>
<td></td>
<td>52,955</td>
<td></td>
</tr>
<tr>
<td>bends</td>
<td></td>
<td>2,018</td>
<td></td>
</tr>
<tr>
<td>Total</td>
<td></td>
<td>87,933</td>
<td>74,899</td>
</tr>
<tr>
<td>30% variation</td>
<td></td>
<td>114,313</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>61,553</td>
<td></td>
</tr>
<tr>
<td>Compressor</td>
<td></td>
<td>656,128</td>
<td>1,126,836</td>
</tr>
<tr>
<td>30% variation</td>
<td></td>
<td>852,967</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>459,290</td>
<td></td>
</tr>
</tbody>
</table>

From the table above, the cost of the reactor with 30% accuracy is in agreement with the reactor cost calculated with six-tenth rule. The price of the compressor calculated from the data of Dautzenberg (2003) is cheaper than the cost calculated with six-tenth rule, so the six-tenth rule can provide easy and rapid cost calculation. This method is used to calculate the purchase equipment cost for the plant capacities of 80,000 and 8,000 ton/year.
## 11. Operating cost calculation

### 1. Operating cost and required selling price for 125,000 ton/year in the Netherlands

#### Fixed Capital
- **US $ 16,218,193**

#### Working Capital
- **US $ 3,582,825**

#### Start Up Cost
- **US $ 10,748,475**

#### TOTAL CAPITAL COST
- **US $ 30,549,493**

<table>
<thead>
<tr>
<th>location</th>
<th>VARIABLE COST</th>
<th>Plant Capacity 125,000MT / year</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>US $</td>
<td>US $ / T</td>
</tr>
<tr>
<td><strong>Raw Material</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Waste cooking oil</td>
<td>388US$/T</td>
<td>0.99T / MT</td>
<td>48,175,498</td>
<td>385.40</td>
</tr>
<tr>
<td>Methanol</td>
<td>300US$/T</td>
<td>0.11T / MT</td>
<td>4,218,750</td>
<td>33.75</td>
</tr>
<tr>
<td>price of waste cooking oil = 0.30 euro/liter</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Total Raw Materials</strong></td>
<td></td>
<td></td>
<td><strong>52,394,248</strong></td>
<td><strong>33.75</strong></td>
</tr>
</tbody>
</table>

| **Start Up / Recycling afterwards** | | | | |
| Methanol | 300US$/T | 48Ton | 14,400 | 0.12 |
| Propane | 1102US$/T | 4Ton | 4,409 | 0.04 |
| **Total Start-up** | | | **18,809** | **0.15** |

| **Utilities** | | | | |
| Electricity | 7.9E-02US$/kWh | 335.68kWh / MT | 3,299,760 | 26.40 |
| Cooling water | 1.3US$/T | 0.96T / MT | 150,000 | 1.2E+00 |
| **Total Utilities** | | | **3,449,760** | **27.60** |

| **By-product Credit** | | | | |
| Glycerol | 1200US $/T | 0.11T / MT | 15,937,500 | 127.50 |
| price of 92% pure glycerine | | | | |
| **Total By-product Credit** | | | **15,937,500** | **127.50** |

| **Fixed Cost** | | | | |
| Operating Labor | 3operators shift | 240k$/year | 720,000 | 5.76 |
| 1supervisors shift | 300k$/year | 300,000 | 2.40 |
| Maintenance | 5 % of ISBL + 3 % of OSBL | | 756,849 | 6.05 |
| Plant Overhead | 80 % of operating labor + 20 % of maintenance | | 967,370 | 7.74 |
| Taxes & Insurance | 2% of fixed capital | | 324,364 | 2.59 |
| **Total Fixed Cost** | | | **3,068,583** | **24.55** |

| **TOTAL OPERATING COST** | | | **42,993,901** | **-41.45** |

| Capital Charges | 20 % DCF-ROI (2 years construction) | | 9,381,533 | 75.05 |
| S, G & A | 5% of RSP | | 2,618,772 | 20.95 |
| **REQUIRED SELLING PRICE** | | | **54,994,205** | **439.95** |

| RSP (US $ / TON) | 440 |
| RSP (US $ / KG) | 0.440 |
| RSP (US $ / liter) | 0.37 |
2. Operating cost and required selling price for 125,000 ton/year in United States

<table>
<thead>
<tr>
<th>Location</th>
<th>United States</th>
<th>Plant Capacity</th>
<th>125,000MT / year</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>VARIABLE COST</th>
<th>cost/unit</th>
<th>consumption factor</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Raw Material</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Waste cooking oil</td>
<td>210US$/T</td>
<td>0.99T / MT</td>
<td>26,068,993</td>
<td>208.55</td>
</tr>
<tr>
<td>Methanol</td>
<td>300US$/T</td>
<td>0.11T / MT</td>
<td>4,218,750</td>
<td>33.75</td>
</tr>
<tr>
<td><strong>Total Raw Materials</strong></td>
<td></td>
<td></td>
<td>30,287,743</td>
<td>242.30</td>
</tr>
</tbody>
</table>

| **Start Up / Recycling afterwards** | | | | |
| Methanol | 300US$/T | 48Ton | 14,400 | 0.12 |
| Propane  | 1102US$/T| 4Ton  | 4,409  | 0.04 |
| **Total Start-up** | | | 18,809 | 0.15 |

<table>
<thead>
<tr>
<th><strong>Utilities</strong></th>
<th>cost/energy</th>
<th>consumption factor</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Electricity</td>
<td>8.0E-02US$/kWh</td>
<td>335.68kWh / MT</td>
<td>3,335,820</td>
<td>26.69</td>
</tr>
<tr>
<td>Cooling water</td>
<td>8.6E-01US$/T</td>
<td>0.96T / MT</td>
<td>102,708</td>
<td>8.2E-01</td>
</tr>
<tr>
<td><strong>Total Utilities</strong></td>
<td></td>
<td></td>
<td>3,438,528</td>
<td>27.51</td>
</tr>
</tbody>
</table>

| **By-product Credit** | cost/unit | consumption factor | Annual cost | Unit cost |
| Glycerol          | 1200US $/T | 0.11T / MT         | 15,937,500 | 127.50 |
| **Total By-product Credit** | | | 15,937,500 | 127.50 |

| **Fixed Cost** | cost/year | percentage of capital | Annual cost | Unit cost |
| Operating Labor | 3operators shift | 240k$/year | 720,000 | 5.76 |
| Maintenance     | 1supervisors shift | 300k$/year | 756,849 | 6.05 |
| Plant Overhead  | 5 % of ISBL + 3 % of OSBL | 967,370 | 7.74 |
| Taxes & Insurance | 80 % of operating labor + 20 % of maintenance | 324,364 | 2.59 |
| **Total Fixed Cost** | | | 3,068,583 | 24.55 |

**TOTAL OPERATING COST** 20,876,163 167.01

**REQUIRED SELLING PRICE** 29,734,643 237.88

| RSP (US $ / TON) | 238 |
| RSP (US $ / KG)  | 0.238 |
| RSP (US $ / liter) | 0.20 |
3. Operating cost and required selling price for 125,000 ton/year in US with bioethanol

<table>
<thead>
<tr>
<th>Location</th>
<th>United States</th>
<th>Plant Capacity</th>
<th>125,000MT / year</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Fixed Capital</strong></td>
<td>US $ 17,392,801</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Working Capital</strong></td>
<td>US $ 1,811,651</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Start Up Cost</strong></td>
<td>US $ 5,434,954</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>TOTAL CAPITAL COST</strong></td>
<td>US $ 24,639,406</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Raw Material</th>
<th>cost/unit</th>
<th>consumption factor</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Waste cooking oil</td>
<td>210US$/T</td>
<td>0.99T / MT</td>
<td>26,068,993</td>
<td>208.55</td>
</tr>
<tr>
<td>Bioethanol</td>
<td>355US$/T</td>
<td>0.11T / MT</td>
<td>4,990,494</td>
<td>39.92</td>
</tr>
<tr>
<td><strong>Total Raw Materials</strong></td>
<td></td>
<td></td>
<td>31,059,487</td>
<td>248.48</td>
</tr>
</tbody>
</table>

| Start Up / Recycling afterwards |          |                   |             |           |
| Ethanol                        | 355US$/T  | 48Ton             | 17,040      | 0.14      |
| Propane                        | 1102US$/T | 4Ton              | 4,409       | 0.04      |
| **Total Start-up**             |           |                    | 21,449      | 0.17      |

| Utilities                     |           |                   |             |           |
| Electricity                   | 8.0E-02US$/kWh | 335.68kWh / MT | 3,335,820   | 26.69     |
| Cooling water                 | 8.6E-01US$/T | 0.96T / MT       | 102,708     | 8.2E-01   |
| **Total Utilities**           |           |                    | 3,438,528   | 27.51     |

| By-product Credit             |           |                   |             |           |
| Glycerol                      | 1200US $/T | 0.11T / MT        | 15,937,500  | 127.50    |
| **Total By-product Credit**   |           |                    | 15,937,500  | 127.50    |

| Fixed Cost                    |           |                   |             |           |
| Operating Labor               | 3 operators shift 240 k$/year | 720,000     | 5.76      |
| Maintenance                   | 1 supervisors shift 300 k$/year | 300,000     | 2.40      |
| Maintenance                   | 5 % of ISBL + 3 % of OSBL | 811,664     | 6.49      |
| Plant Overhead                | 80 % of operating labor + 20 % of maintenance | 978,333     | 7.83      |
| Taxes & Insurance             | 2% of fixed capital | 347,856     | 2.78      |
| **Total Fixed Cost**          |           |                    | 3,157,853   | 25.26     |

| TOTAL OPERATING COST          |          |                   | 21,739,818  | 173.92    |

| Capital Charges               | 20 % DCF-ROI (2 years construction) | 7,910,643 | 63.29 |
| S, G & A                      | 5% of RSP | 1,482,523 | 11.86 |
| **REQUIRED SELLING PRICE**    |          |                   | 31,132,983  | 249.06    |
| RSP (US $ / TON)              |          |                   | 249         |           |
| RSP (US $ / KG)               |          |                   | 0.249       |           |
| RSP (US $ / liter)            |          |                   | 0.21        |           |
4. Operating cost and required selling price for 80,000 ton/year in the Netherlands

<table>
<thead>
<tr>
<th>Fixed Capital</th>
<th>US $</th>
<th>12,408,247</th>
</tr>
</thead>
<tbody>
<tr>
<td>Working Capital</td>
<td>US $</td>
<td>2,707,373</td>
</tr>
<tr>
<td>Start Up Cost</td>
<td>US $</td>
<td>8,122,119</td>
</tr>
<tr>
<td><strong>TOTAL CAPITAL COST</strong></td>
<td>US $</td>
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<table>
<thead>
<tr>
<th>location</th>
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<tbody>
<tr>
<td><strong>VARIABLE COST</strong></td>
<td>Plant Capacity</td>
</tr>
<tr>
<td></td>
<td>80,000MT / year</td>
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<tr>
<td>cost/unit</td>
<td>Annual cost</td>
</tr>
<tr>
<td></td>
<td>Unit cost</td>
</tr>
<tr>
<td>Raw Material</td>
<td>US $ US$/T</td>
</tr>
<tr>
<td>Waste cooking oil</td>
<td>388US$/T</td>
</tr>
<tr>
<td>Methanol</td>
<td>300US$/T</td>
</tr>
<tr>
<td><strong>Total Raw Materials</strong></td>
<td>33,752,925</td>
</tr>
<tr>
<td></td>
<td>34.20</td>
</tr>
<tr>
<td>Start Up / Recycling afterwards</td>
<td></td>
</tr>
<tr>
<td>Methanol</td>
<td>300US$/T</td>
</tr>
<tr>
<td>Propane</td>
<td>1102US$/T</td>
</tr>
<tr>
<td><strong>Total Start-up</strong></td>
<td>11,722</td>
</tr>
<tr>
<td></td>
<td>0.15</td>
</tr>
<tr>
<td>Utilities</td>
<td>US $ US$/T</td>
</tr>
<tr>
<td>Electricity</td>
<td>0.1US$/kWh</td>
</tr>
<tr>
<td>Cooling water</td>
<td>1.3US$/T</td>
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<tr>
<td><strong>Total Utilities</strong></td>
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<tr>
<td>By-product Credit</td>
<td>US $ US$/T</td>
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<tr>
<td>Glycerol</td>
<td>1200US $/T</td>
</tr>
<tr>
<td><strong>Total By-product Credit</strong></td>
<td>6,234,000</td>
</tr>
<tr>
<td></td>
<td>77.93</td>
</tr>
<tr>
<td>Fixed Cost</td>
<td>US $ US$/T</td>
</tr>
<tr>
<td>Operating Labor</td>
<td>3operators shift</td>
</tr>
<tr>
<td>1supervisors shift</td>
<td>240k$/year</td>
</tr>
<tr>
<td>300k$/year</td>
<td>720,000</td>
</tr>
<tr>
<td>Maintenance</td>
<td>5% of ISBL + 3% of OSBL</td>
</tr>
<tr>
<td>Plant Overhead</td>
<td>80% of operating labor + 20% of maintenance</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
<td>2% of fixed capital</td>
</tr>
<tr>
<td><strong>Total Fixed Cost</strong></td>
<td>2,779,027</td>
</tr>
<tr>
<td></td>
<td>34.74</td>
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</tbody>
</table>

**TOTAL OPERATING COST** 32,488,478 18.39

Capital Charges 20 % DCF-ROI (2 years construction) 7,187,822 89.85
S, G & A 5% of RSP 1,983,815 24.80

**REQUIRED SELLING PRICE** 41,660,115 520.75
RSP (US $ / TON) 521
RSP (US $ / KG) 0.521
RSP (US $ / liter) 0.44
5. Operating cost and required selling price for 80,000 ton/year in United States

### Fixed Capital
- US $12,408,247

### Working Capital
- US $1,520,620

### Start Up Cost
- US $4,561,861

### TOTAL CAPITAL COST
- US $18,490,729

<table>
<thead>
<tr>
<th>Location</th>
<th>United States</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>VARIABLE COST</strong></td>
<td></td>
</tr>
<tr>
<td><strong>Plant Capacity</strong></td>
<td><strong>80,000MT / year</strong></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Raw Material</th>
<th>cost/unit</th>
<th>consumption factor</th>
<th>Annual cost US $</th>
<th>Unit cost US $ / T</th>
</tr>
</thead>
<tbody>
<tr>
<td>Waste cooking oil</td>
<td>210US$/T</td>
<td>1.00T / MT</td>
<td>16,784,050</td>
<td>209.80</td>
</tr>
<tr>
<td>Methanol</td>
<td>300US$/T</td>
<td>0.11T / MT</td>
<td>2,736,000</td>
<td>34.20</td>
</tr>
<tr>
<td><strong>Total Raw Materials</strong></td>
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<td></td>
<td>19,520,050</td>
<td>244.00</td>
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</table>

<table>
<thead>
<tr>
<th>Start Up / Recycling afterwards</th>
<th>cost/unit</th>
<th>Annual cost US $</th>
<th>Unit cost US $ / T</th>
</tr>
</thead>
<tbody>
<tr>
<td>Methanol</td>
<td>300US$/T</td>
<td>9,050</td>
<td>0.11</td>
</tr>
<tr>
<td>Propane</td>
<td>1102US$/T</td>
<td>2,672</td>
<td>0.03</td>
</tr>
<tr>
<td><strong>Total Start-up</strong></td>
<td></td>
<td>11,722</td>
<td>0.15</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Utilities</th>
<th>cost/unit</th>
<th>Annual cost US $</th>
<th>Unit cost US $ / T</th>
</tr>
</thead>
<tbody>
<tr>
<td>Electricity</td>
<td>8.0E-02US$/kWh</td>
<td>330.75kWh / MT</td>
<td>2,103,542</td>
</tr>
<tr>
<td>Cooling water</td>
<td>8.6E-01US$/T</td>
<td>0.98T / MT</td>
<td>67,103</td>
</tr>
<tr>
<td><strong>Total Utilities</strong></td>
<td></td>
<td>2,170,645</td>
<td>27.13</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>By-product Credit</th>
<th>cost/unit</th>
<th>Annual cost US $</th>
<th>Unit cost US $ / T</th>
</tr>
</thead>
<tbody>
<tr>
<td>Glycerol</td>
<td>1200US$/T</td>
<td>6,234,000</td>
<td>77.93</td>
</tr>
<tr>
<td><strong>Total By-product Credit</strong></td>
<td></td>
<td>6,234,000</td>
<td>77.93</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Fixed Cost</th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Operating Labor</td>
<td>3operators shift</td>
<td>240k$/year</td>
<td>720,000</td>
</tr>
<tr>
<td></td>
<td>1supervisors shift</td>
<td>300k$/year</td>
<td>300,000</td>
</tr>
<tr>
<td>Maintenance</td>
<td>5 % of ISBL + 3 % of OSBL</td>
<td>579,052</td>
<td>7.24</td>
</tr>
<tr>
<td>Plant Overhead</td>
<td>80 % of operating labor + 20 % of maintenance</td>
<td>931,810</td>
<td>11.65</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
<td>2% of fixed capital</td>
<td>248,165</td>
<td>3.10</td>
</tr>
<tr>
<td><strong>Total Fixed Cost</strong></td>
<td></td>
<td>2,779,027</td>
<td>34.74</td>
</tr>
</tbody>
</table>

**TOTAL OPERATING COST**: 18,247,445 228.09

<table>
<thead>
<tr>
<th>Capital Charges</th>
<th>20 % DCF-ROI (2years construction)</th>
<th>5% of RSP</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>5,939,358</td>
<td>74.24</td>
</tr>
<tr>
<td></td>
<td>1,209,340</td>
<td>15.12</td>
</tr>
</tbody>
</table>

**REQUIRED SELLING PRICE**
- **RSP (US $ / TON)**: 317
- **RSP (US $ / KG)**: 0.317
- **RSP (US $ / liter)**: 0.27
6. Operating cost and required selling price for 80,000 ton/year in US with bioethanol

<table>
<thead>
<tr>
<th></th>
<th>United States</th>
<th>Plant Capacity</th>
</tr>
</thead>
<tbody>
<tr>
<td>location</td>
<td></td>
<td>80,000MT / year</td>
</tr>
<tr>
<td><strong>VARIABLE COST</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>cost/unit</td>
<td>Annual cost</td>
<td>Unit cost</td>
</tr>
<tr>
<td></td>
<td>consumption factor</td>
<td>US $</td>
</tr>
<tr>
<td><strong>Raw Material</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Waste cooking oil</td>
<td>210US$/T</td>
<td>1.00T / MT</td>
</tr>
<tr>
<td>Bioethanol</td>
<td>355US$/T</td>
<td>0.11T / MT</td>
</tr>
<tr>
<td><strong>Total Raw Materials</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Start Up / Recycling afterwards</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Bio-ethanol</td>
<td>355US$/T</td>
<td>30Ton</td>
</tr>
<tr>
<td>Propane</td>
<td>1102US$/T</td>
<td>2Ton</td>
</tr>
<tr>
<td><strong>Total Start-up</strong></td>
<td>13,382</td>
<td>0.17</td>
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<tr>
<td><strong>Utilities</strong></td>
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<td></td>
</tr>
<tr>
<td>Electricity</td>
<td>8.0E-02US$/kWh</td>
<td>330.75kWh / MT</td>
</tr>
<tr>
<td>Cooling water</td>
<td>8.6E-01US$/T</td>
<td>0.98T / MT</td>
</tr>
<tr>
<td><strong>Total Utilities</strong></td>
<td>2,170,645</td>
<td>27.13</td>
</tr>
<tr>
<td><strong>By-product Credit</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Glycerol</td>
<td>1200US $/T</td>
<td>0.06T / MT</td>
</tr>
<tr>
<td><strong>Total By-product Credit</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Fixed Cost</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Operating Labor</td>
<td>240k$/year</td>
<td>720,000</td>
</tr>
<tr>
<td>Maintenance</td>
<td>5 % of ISBL + 3 % of OSBL</td>
<td>620,990</td>
</tr>
<tr>
<td>Plant Overhead</td>
<td>80 % of operating labor + 20 % of maintenance</td>
<td>940,198</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
<td>2% of fixed capital</td>
<td>266,138</td>
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<tr>
<td><strong>Total Fixed Cost</strong></td>
<td>2,847,326</td>
<td>35.59</td>
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<tr>
<td><strong>TOTAL OPERATING COST</strong></td>
<td>18,817,905</td>
<td>235.22</td>
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</table>

**REQUIRED SELLING PRICE**

- RSP (US $ / TON): 26,362,854 / 330
- RSP (US $ / KG): 0.330
- RSP (US $ / liter): 0.28
7. Operating cost and required selling price for 8,000 ton/year in United States (20% ROI)

<table>
<thead>
<tr>
<th>Location</th>
<th>Fixed Capital</th>
<th>Working Capital</th>
<th>Start Up Cost</th>
<th>TOTAL CAPITAL COST</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>VARIABLE COST</th>
<th>location</th>
<th>Plant Capacity</th>
<th>8,000MT / year</th>
</tr>
</thead>
<tbody>
<tr>
<td>cost/unit</td>
<td>consumption factor</td>
<td>Annual cost</td>
<td>Unit cost</td>
</tr>
<tr>
<td>Raw Material</td>
<td>Waste cooking oil</td>
<td>210US$/T</td>
<td>1.02T / MT</td>
</tr>
<tr>
<td>Methanol</td>
<td>300US$/T</td>
<td>0.12T / MT</td>
<td>US $ 278,400</td>
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<td>Total Raw Materials</td>
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<td>US $ 1,987,529</td>
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<tr>
<td>Start Up / Recycling afterwards</td>
<td>Methanol</td>
<td>300US$/T</td>
<td>3Ton</td>
</tr>
<tr>
<td>Propane</td>
<td>1102US$/T</td>
<td>0.2Ton</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Total Start-up</td>
<td></td>
<td>US $ 1,193</td>
</tr>
<tr>
<td>Utilities</td>
<td>Electricity</td>
<td>8.0E-02US$/kWh</td>
<td>356.00kWh / MT</td>
</tr>
<tr>
<td>Cooling water</td>
<td>8.6E-01US$/T</td>
<td>1.20T / MT</td>
<td>US $ 8,217</td>
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<tr>
<td></td>
<td>Total Utilities</td>
<td></td>
<td>US $ 234,633</td>
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<td>By-product Credit</td>
<td>Glycerol</td>
<td>1200US $/T</td>
<td>0.11T / MT</td>
</tr>
<tr>
<td>Fixed Cost</td>
<td>Operating Labor</td>
<td>3operators shift</td>
<td>240k$/year</td>
</tr>
<tr>
<td></td>
<td>1supervisors shift</td>
<td>300k$/year</td>
<td></td>
</tr>
<tr>
<td>Maintenance</td>
<td>5 % of ISBL + 3 % of OSBL</td>
<td></td>
<td>US $ 579,052</td>
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<tr>
<td>Plant Overhead</td>
<td>80 % of operating labor + 20 % of maintenance</td>
<td></td>
<td>US $ 931,810</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
<td>2% of fixed capital</td>
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<td>US $ 62,336</td>
</tr>
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<td></td>
<td>Total Fixed Cost</td>
<td></td>
<td>US $ 2,593,198</td>
</tr>
<tr>
<td></td>
<td>TOTAL OPERATING COST</td>
<td></td>
<td>US $ 3,798,953</td>
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<tr>
<td>Capital Charges</td>
<td>20 % DCF-ROI (2 years construction)</td>
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<td>US $ 1,624,001</td>
</tr>
<tr>
<td>S, G &amp; A</td>
<td>5% of RSP</td>
<td></td>
<td>US $ 271,148</td>
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<tr>
<td>REQUIRED SELLING PRICE</td>
<td>RSP (US $ / TON)</td>
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<td>US $ 5,694,101</td>
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<tr>
<td></td>
<td>RSP (US $ / KG)</td>
<td></td>
<td>US $ 0.712</td>
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<tr>
<td></td>
<td>RSP (US $ / liter)</td>
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<td>US $ 0.60</td>
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</table>
8. Operating cost and required selling price for 8,000 ton/year in United States (15% ROI) to compare with result of Zhang et al (2003)

<table>
<thead>
<tr>
<th>Fixed Capital</th>
<th>US $  3,116,811</th>
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</thead>
<tbody>
<tr>
<td>Working Capital</td>
<td>US $  316,579</td>
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<tr>
<td>Start Up Cost</td>
<td>US $  949,738</td>
</tr>
<tr>
<td>TOTAL CAPITAL COST</td>
<td>US $  4,383,128</td>
</tr>
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</table>

<table>
<thead>
<tr>
<th>location</th>
<th>United States</th>
<th>Plant Capacity 8,000MT / year</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>VARIABLE COST</th>
<th>Annual cost</th>
<th>Unit cost</th>
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<tbody>
<tr>
<td></td>
<td>US $</td>
<td>US $ / T</td>
</tr>
<tr>
<td>cost/unit</td>
<td>consumption factor</td>
<td></td>
</tr>
<tr>
<td>Raw Material</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Waste cooking oil</td>
<td>210US$/T</td>
<td>1.02T / MT</td>
</tr>
<tr>
<td>Methanol</td>
<td>300US$/T</td>
<td>0.12T / MT</td>
</tr>
<tr>
<td>Total Raw Materials</td>
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<td></td>
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<table>
<thead>
<tr>
<th>Start Up / Recycling afterwards</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>US $</td>
<td>US $ / T</td>
</tr>
<tr>
<td>Methanol</td>
<td>300US$/T</td>
<td>3T on</td>
</tr>
<tr>
<td>Propane</td>
<td>1102US$/T</td>
<td>0.2T on</td>
</tr>
<tr>
<td>Total Start-up</td>
<td>1,193</td>
<td>0.15</td>
</tr>
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</table>

<table>
<thead>
<tr>
<th>Utilities</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>US $</td>
<td>US $ / T</td>
</tr>
<tr>
<td>Electricity</td>
<td>8.0E-02US$/kWh</td>
<td>356.00kWh / MT</td>
</tr>
<tr>
<td>Cooling water</td>
<td>8.6E-01US$/T</td>
<td>1.20T / MT</td>
</tr>
<tr>
<td>Total Utilities</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>By-product Credit</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>US $</td>
<td>US $ / T</td>
</tr>
<tr>
<td>Glycerol</td>
<td>1200US$/T</td>
<td>0.11T / MT</td>
</tr>
<tr>
<td>Total By-product Credit</td>
<td>1,017,600</td>
<td>127.20</td>
</tr>
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</table>

<table>
<thead>
<tr>
<th>Fixed Cost</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>US $</td>
<td>US $ / T</td>
</tr>
<tr>
<td>Operating Labor</td>
<td>3operators shift</td>
<td>240k$/year</td>
</tr>
<tr>
<td></td>
<td>1supervisors shift</td>
<td>300k$/year</td>
</tr>
<tr>
<td>Maintenance</td>
<td>5% of ISBL + 3% of OSBL</td>
<td>579,052</td>
</tr>
<tr>
<td>Plant Overhead</td>
<td>80% of operating labor + 20% of maintenance</td>
<td>931,810</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
<td>2% of fixed capital</td>
<td>62,336</td>
</tr>
<tr>
<td>Total Fixed Cost</td>
<td>2,593,198</td>
<td>324.15</td>
</tr>
</tbody>
</table>

TOTAL OPERATING COST 3,798,953 474.87

<table>
<thead>
<tr>
<th>Capital Charges</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>US $</td>
<td>US $ / T</td>
</tr>
<tr>
<td>15% DCF-ROI (2 years construction)</td>
<td>1,225,056</td>
<td>153.13</td>
</tr>
<tr>
<td>5% of RSP</td>
<td>251,200</td>
<td>31.40</td>
</tr>
</tbody>
</table>

REQUIRED SELLING PRICE
RSP (US $ / TON) 659
RSP (US $ / KG) 0.659
RSP (US $ / liter) 0.55
9. Operating cost and required selling price for 8,000 ton/year in the Netherlands

<table>
<thead>
<tr>
<th></th>
<th>Fixed Capital</th>
<th>Working Capital</th>
<th>Start Up Cost</th>
<th>TOTAL CAPITAL COST</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>US $</td>
<td>US $</td>
<td>US $</td>
<td>US $</td>
</tr>
<tr>
<td>Value</td>
<td>3,116,811</td>
<td>437,358</td>
<td>1,312,074</td>
<td>4,866,242</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Location</th>
<th>Netherlands</th>
<th>Plant Capacity</th>
</tr>
</thead>
<tbody>
<tr>
<td>Value</td>
<td></td>
<td>8,000MT / year</td>
</tr>
</tbody>
</table>

**VARIABLE COST**

<table>
<thead>
<tr>
<th>Raw Material</th>
<th>cost/unit</th>
<th>consumption factor</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Waste cooking oil</td>
<td>388US$/T</td>
<td>1.02T / MT</td>
<td>3,158,471</td>
<td>394.81</td>
</tr>
<tr>
<td>Methanol</td>
<td>300US$/T</td>
<td>0.12T / MT</td>
<td>278,400</td>
<td>34.80</td>
</tr>
<tr>
<td><strong>Total Raw Materials</strong></td>
<td></td>
<td></td>
<td>3,436,871</td>
<td>429.61</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Start Up / Recycling afterwards</th>
</tr>
</thead>
<tbody>
<tr>
<td>Methanol</td>
</tr>
<tr>
<td>Propane</td>
</tr>
<tr>
<td><strong>Total Start-up</strong></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Utilities</th>
<th>US$/kW</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Electricity</td>
<td>8.0E-02h</td>
<td>356.00kWh / MT</td>
<td>226,416</td>
</tr>
<tr>
<td>Cooling water</td>
<td>8.6E-01US$/T</td>
<td>1.20T / MT</td>
<td>8,217</td>
</tr>
<tr>
<td><strong>Total Utilities</strong></td>
<td></td>
<td></td>
<td>234,633</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>By-product Credit</th>
<th>cost/unit</th>
<th>consumption factor</th>
<th>Annual cost</th>
<th>Unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Glycerol</td>
<td>1200US$/T</td>
<td>0.11T / MT</td>
<td>1,017,600</td>
<td>127.20</td>
</tr>
<tr>
<td><strong>Total By-product Credit</strong></td>
<td></td>
<td></td>
<td>1,017,600</td>
<td>127.20</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Fixed Cost</th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Operating Labor</td>
<td>240k$/year</td>
<td>720,000</td>
<td>90.00</td>
<td></td>
</tr>
<tr>
<td>Maintenance</td>
<td>300k$/year</td>
<td>579,052</td>
<td>72.38</td>
<td></td>
</tr>
<tr>
<td>Plant Overhead</td>
<td>931,810</td>
<td>116.48</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
<td>62,336</td>
<td>7.79</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Total Fixed Cost</strong></td>
<td></td>
<td></td>
<td>2,593,198</td>
<td>324.15</td>
</tr>
</tbody>
</table>

**TOTAL OPERATING COST**

|                  |               |                   |             |           |
|                  | 5,248,294     | 656.04            |

| Capital Charges   | 20 % DCF-ROI  | ( 2 years construction ) | 1,751,060   | 218.88    |
| S, G & A          | 5% of RSP     |                         | 349,968     | 43.75     |

<table>
<thead>
<tr>
<th>REQUIRED SELLING PRICE</th>
</tr>
</thead>
<tbody>
<tr>
<td>RSP (US $ / TON)</td>
</tr>
<tr>
<td>RSP (US $ / KG)</td>
</tr>
<tr>
<td>RSP (US $ / liter)</td>
</tr>
</tbody>
</table>
10. Operating cost and required selling price for 8,000 ton/year in US with bioethanol

<table>
<thead>
<tr>
<th>Fixed Capital</th>
<th>US $3,342,547</th>
</tr>
</thead>
<tbody>
<tr>
<td>Working Capital</td>
<td>US $321,214</td>
</tr>
<tr>
<td>Start Up Cost</td>
<td>US $963,641</td>
</tr>
<tr>
<td>TOTAL CAPITAL COST</td>
<td>US $4,627,402</td>
</tr>
</tbody>
</table>

| location                  | United States | Plant Capacity          | 8,000MT / year |
|----------------------------|----------------|-------------------------|

**VARIABLE COST**

| cost/unit                  | consumption factor | Annual cost | Unit cost  
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Raw Material</td>
<td></td>
<td>US $</td>
<td>US $ / T</td>
</tr>
<tr>
<td>Waste cooking oil</td>
<td>210US$/T</td>
<td>1.02T / MT</td>
<td>1,709,129</td>
</tr>
<tr>
<td>Bioethanol</td>
<td>355US$/T</td>
<td>0.12T / MT</td>
<td>329,328</td>
</tr>
<tr>
<td>Total Raw Materials</td>
<td></td>
<td></td>
<td>2,038,457</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Start Up / Recycling afterwards</th>
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<tbody>
<tr>
<td>Bioethanol</td>
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<tr>
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<table>
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<tr>
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<tbody>
<tr>
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<tr>
<td>Cooling water</td>
</tr>
<tr>
<td>Total Utilities</td>
</tr>
</tbody>
</table>

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<tr>
<th>By-product Credit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Glycerol</td>
</tr>
<tr>
<td>Total By-product Credit</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Fixed Cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Operating Labor</td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td>Maintenance</td>
</tr>
<tr>
<td>Plant Overhead</td>
</tr>
<tr>
<td>Taxes &amp; Insurance</td>
</tr>
<tr>
<td>Total Fixed Cost</td>
</tr>
</tbody>
</table>

**TOTAL OPERATING COST**

<table>
<thead>
<tr>
<th>Capital Charges</th>
</tr>
</thead>
<tbody>
<tr>
<td>20 % DCF-ROI ( 2 years construction )</td>
</tr>
<tr>
<td>5% of RSP</td>
</tr>
<tr>
<td>REQUIRED SELLING PRICE</td>
</tr>
<tr>
<td>RSP (US $ / TON)</td>
</tr>
<tr>
<td>RSP (US $ / KG)</td>
</tr>
<tr>
<td>RSP (US $ / liter)</td>
</tr>
</tbody>
</table>

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SUMMARY

The world is facing a constant decrease of energy supply especially from fossil derived sources. One of the alternatives which is already produced small scale is biodiesel.

Biodiesel can be produced from waste cooking oil. Since waste cooking oil contains water and free fatty acids, supercritical transesterification offers great advantage to decrease the capital and manufacturing cost of the biodiesel.

The supercritical transesterification for biodiesel continuous production have been studied technically and economically. It can be concluded that biodiesel by supercritical transesterification is technically feasible with high purity of methyl esters. Almost pure glycerol also attained as by-product.

The economic assessment of the biodiesel plant has been carried out. It can be concluded that the biodiesel is economically feasible because the price is lower than catalysed production biodiesel and also lower than normal fossil based diesel.

The sensitive key factors for the economic feasibility of the plant are: raw material price, plant capacity, glycerol price and capital cost.

Keywords: biodiesel, supercritical transesterification, process design, economic feasibility