

# Techno-economic analysis of processes for biodiesel production with integrated co-production of higher added value products from glycerol

Luma Sh. Al-Saadi<sup>1a</sup>, Valentine C. Eze<sup>1</sup>, Adam P. Harvey<sup>1</sup>

<sup>1</sup>School of Engineering, Newcastle University  
Newcastle upon Tyne, NE1 7RU, United Kingdom.

<sup>a</sup>Corresponding author: [lumashihab201@gmail.com](mailto:lumashihab201@gmail.com)

## Abstract

A techno-economic analysis was conducted to compare 3 different processes for biodiesel production: (i) the conventional base-catalysed process (NaOH) (ii) a reactive coupling process to form glycerol carbonate (GLC) as a co-product *in situ* (iii) a reactive coupling to form solketal as a co-product *in situ*. The processes were simulated at rapeseed oil feed of 100,000 t/y. The capital investment and energy consumption of biodiesel production with associated GLC formation were found to be lower than that for the conventional process. The GLC process required \$7.63M capital investment and 2.2MW per annum. The conventional process required \$8.75M capital investment and 5.4 MW per annum. Solketal production, however, was more capital-intensive, with \$12.87M capital investment and 25.8MW. The 20-year net present values (NPVs) for the three biodiesel processes were: \$65 M for the conventional alkali-biodiesel process, \$128 M for solketal co-production and \$631 M for GLC co-production. Clearly, the conversion of glycerol into higher added value species *in situ* can significantly increase the profitability of biodiesel production, particularly for glycerol carbonate formation.

**Keywords:** techno-economic analysis, biodiesel production, reactive coupling, glycerol carbonate, solketal.

## 1 Introduction

Due to its renewability and environment benefits, biodiesel has received significant attention as an alternative fuel to petroleum diesel over the last 25y [1-3]. Transesterification is the most commonly used method in commercial biodiesel production in industry, involving reaction of triglyceride (oils or fats) with alcohols, such as methanol (MeOH) in the presence of catalysts. The most common catalysts are base catalysts, particularly sodium methoxide (NaOMe), due to its high rate of reaction. The reaction produces fatty acid alkyl ester (biodiesel) as a main product, and crude glycerol as a by-product [4, 5]. Crude glycerol typically consists of 10%

(wt/wt) of the biodiesel produced, but commands a low market price [6-8]. For this reason, a variety of new biodiesel processing methods have been developed to eliminate or substantially reduce crude glycerol co-production. Use of acetone as a co-reactant for co-production of biodiesel and solketal has been suggested [9]. Similarly, dimethyl carbonate (DMC) has been used as a co-reactant with and without methanol to obtain glycerol carbonate (GLC), rather than crude glycerol [10, 11]. However, for these alternative processes to be implemented for commercial biodiesel production techno-economic analyses are required to assess their profitability. Currently, there are no existing techno-economic analysis data on these alternative processes.

Techno-economic analysis is a technique for investigating and comparing the economic viability and profitability of processes and process technologies. There are techno-economic analyses of various biodiesel production methods [12-16]. One study focussed on the effect of one process step: the purification of crude glycerol from 63.2 to 80 wt.% in first scenario and to 95wt.% in second scenario in a distillation column, at various reflux ratios, from 0.68 to 2.59, corresponding to heat duties of 10.5 kW and 27 kW, respectively [14]. The annual revenues were € 9.32 million and €11.0 million, and the net present values (NPV) were €7.99 million and €9.95 million for each scenario respectively [14], indicating that both were economically viable, but that the second process was preferred.

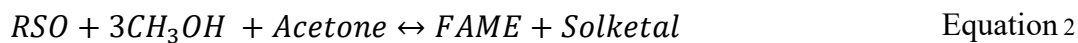
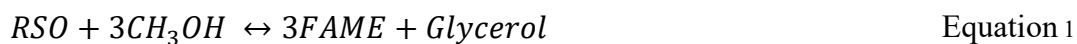
Techno-economic analysis has been used to assess biodiesel production from different processing routes and feedstocks using alkali catalyst: from refined vegetable oil and from pre-treated waste vegetable oil using acid catalyst. In this case, the plant that used waste oil was reported to be the more profitable method as it had significantly reduced manufacturing costs with NPV (9.2) compare with NPV(-26.7) of refined vegetable oil [15].

Another techno-economic study determined the effect of plant production capacity on the profitability. The economic analysis showed that the plants with production capacities in the range 50000-80000 t/y were more economically feasible than those below 15000 t/y [16]. These studies show that technoeconomic analysis is a suitable technique for evaluating competing biodiesel production strategies. Hence, this study uses technoeconomic analyses to investigate the economic viability of biodiesel-solketal and biodiesel-GLC processes, as alternative methods for biodiesel production.

## 2 Materials and process simulation

### 2.1 Materials and chemical reactions

The following chemicals in the Aspen HYSYS database were used: triolein, methyl oleate, sodium hydroxide, glycerol, water, acetic acid, and acetone. The triolein and methyl oleate were selected to be model compounds representing triglycerides from rapeseed oil (RSO) and fatty acid methyl esters (FAME), respectively. The other chemicals used in the process simulations, such as the dimethyl carbonate, guanidine–TBD catalyst, 4-dodecylsulfonic acid catalyst (DBSA) and solketal were estimated using their physical properties via Aspen's user defined compound wizard. Equation 1 was used to describe the chemical reaction and reaction stoichiometry for the conventional biodiesel reaction (alkali-biodiesel), while Equation 2 was used to describe the reactive coupling of biodiesel and solketal (biodiesel-solketal) and Equation 3 the reactive coupling of biodiesel and glycerol carbonate (biodiesel-GLC).



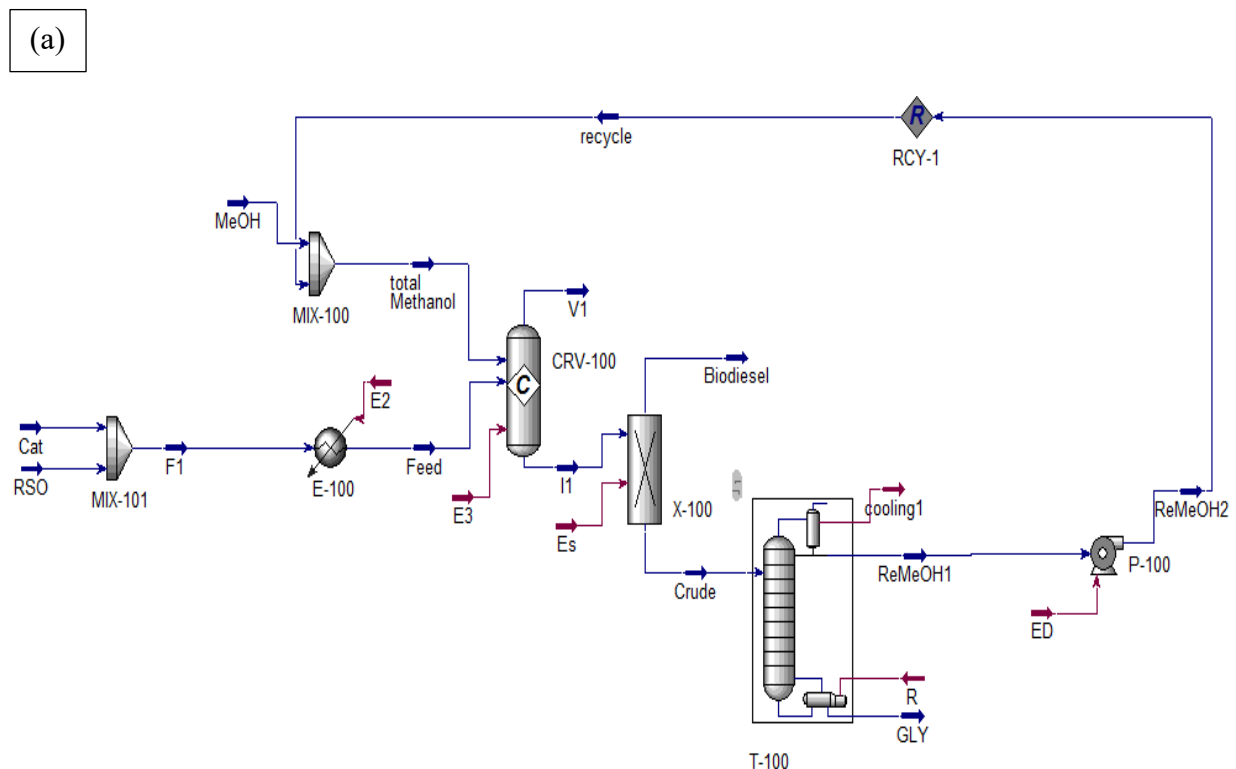
### 2.2 Process design and simulation

The non-random two liquid model was chosen as the thermodynamic model for all three biodiesel processes, as this model is the most suitable for liquid-liquid reactions [17] and polar compounds (such as methanol and glycerol) [15]. The main equipment units including pumps, heat exchangers, cooler, reactors, separators, distillation column were sized and simulated based on Aspen HYSYS data. The conventional alkali-biodiesel process using NaOH catalyst, and the alternative processes for biodiesel-solketal and biodiesel-GLC productions were simulated as continuous processes.

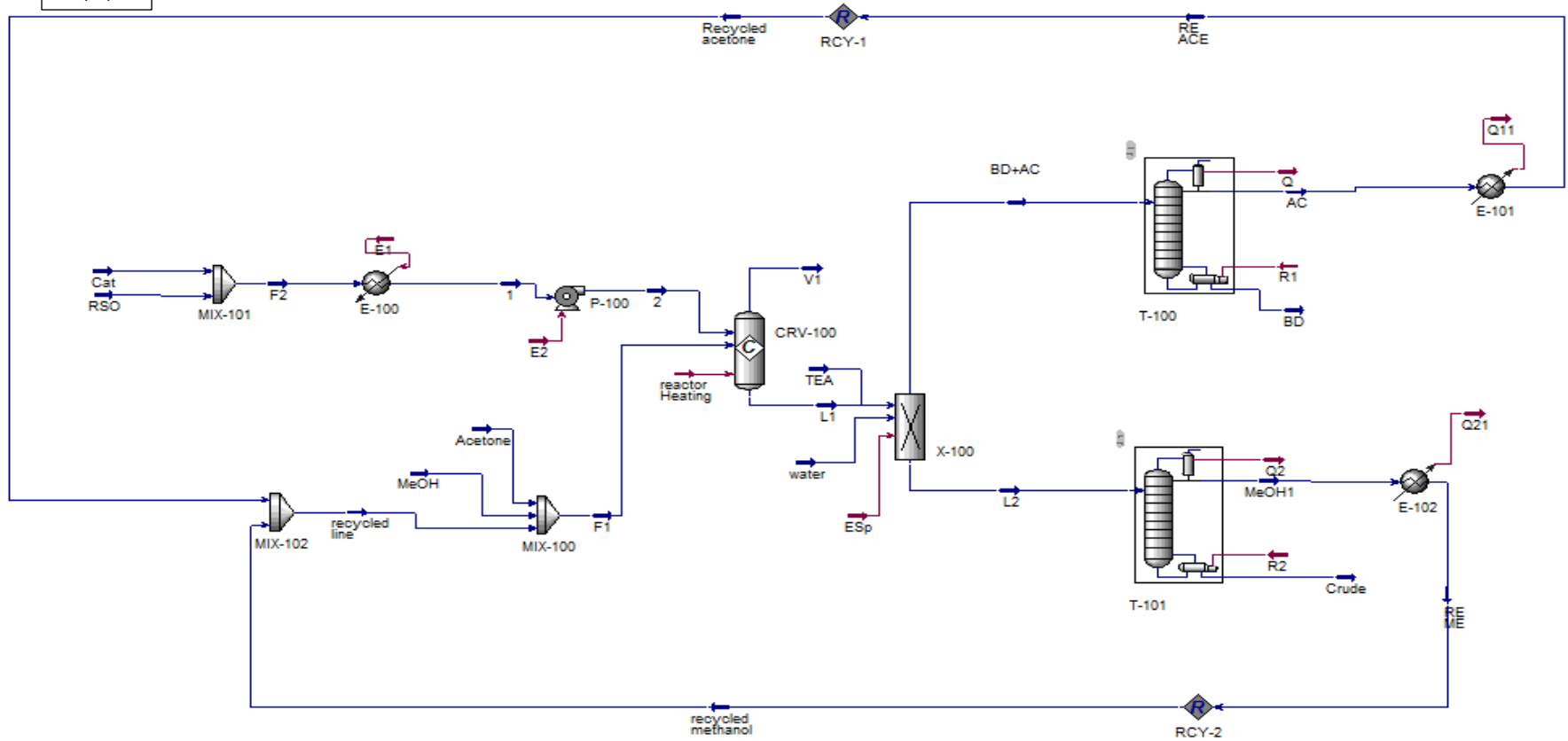
Different operating conditions were required for the different processes, so the basis for comparison was set to be a 98% conversion of the triglyceride to biodiesel, together with the conditions that have been found to maximise co-product yield with that constraint. In the alkali-catalysed biodiesel process, shown in *Figure 1* (a), the conversion of RSO to FAME was set at 98% at 1:6 of RSO-to-methanol molar ratio and 1h, 0.5 wt. % NaOH (based on the oil) and 60

°C [18]. The biodiesel-solketal process was simulated (*Figure 1 (b),*) at a 7:1 acetone-to-triolein (RSO) molar ratio, 10:1 of methanol to oil molar ratio, 50 °C and 0.5 mol of DBSA catalyst to oil, whereas the RSO conversions to FAME was set at 98%, with 82% conversions of the glycerol by-product to solketal after 1h [19]. For the biodiesel-GLC process, the conditions were 2:1 methanol-to-RSO molar ratio, 3:1 DMC-to-RSO molar ratio, 5 wt. % of a TBD-guanidine catalyst and 60 °C [20], as shown in the process flowsheet in *Figure 1(c),* with RSO conversions to FAME and GLC of 98% and 92%, respectively after 4h.

The reactants were preheated to reaction temperature using preheater (E-100) to the reactor (CRV-100). In the diagrams below, the outlet streams from the reactor are the V1 (vent) and the L1 stream, which contains biodiesel, unreacted RSO, unreacted methanol, unreacted acetone, catalyst, glycerol, water, and solketal, DMC, GLC, depending on the plant. The L1 flows to the separator (X-100) to be split into the upper, organic fraction (the “biodiesel-rich phase”) and the lower more polar fraction (glycerol, catalyst, solketal, glycerol carbonate, unreacted methanol, unreacted acetone and water). This stream goes to the distillation column, where the unreacted methanol, in the Alkali-BD and GLC-BD plants, and methanol, or the unreacted acetone, in the Solketal-BD plant, are separated from crude glycerol, catalyst, solketal and GLC compounds. The methanol and acetone are recycled.



(b)



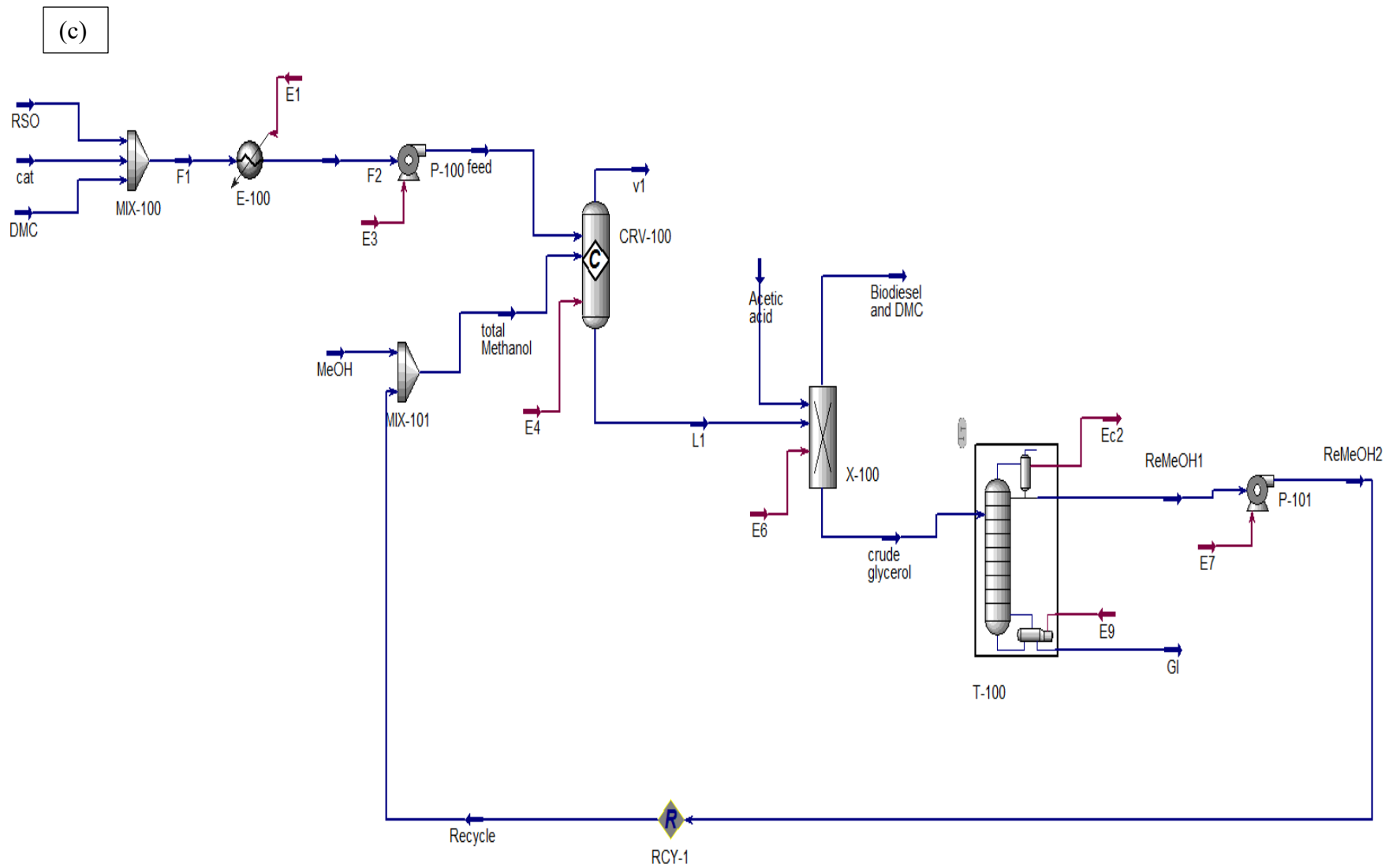


Figure 1: process flowsheets for the Aspen HYSYS simulations (a) alkali-biodiesel process, (b) biodiesel-solketal process, (c) biodiesel-GLC process

### **2.3 Economic analysis**

The economic analysis of the various biodiesel processes in this study were based on the following assumptions:

- I. RSO feed of 100,000 t/y
- II. Plant lifetime of 20 y
- III. Pump efficiency of 75%.
- IV. Equipment purchase costs from the HYSYS database
- V. The total investment cost was calculated based on the investment cost that required to build the plant in addition to operating cost
- VI. The biodiesel downstream purification train and water washing steps in all 3 plants are broadly the same in terms of cost. Based on previous studies they are 6.4% and 5.2 % of the total capital investment respectively [21].

## **3 Results and discussion**

### **3.1 Process simulation**

The material balance for the conventional and alternative routes are given in *Table 1*. The input flow rate of rapeseed oil was 12.2 kmol/h for all the processes. 98% RSO conversion to fatty acid methyl esters was specified for each process, resulting in FAME productions at 35.1 kmol/h. The total flowrates of the “biodiesel-rich phase” from the separator were 35.1 kmol/h for the alkali-biodiesel conventional process, 52.1 kmol/h for the biodiesel-GLC process, and 38.1 kmol/h for the biodiesel-solketal process. The higher flowrates of the biodiesel phases for the biodiesel-GLC and the biodiesel-solketal were due to the miscibility of the unreacted dimethyl carbonate and the acetone (respectively) with the FAME. The biodiesel-GLC process may not require total removal of the unreacted dimethyl carbonate or acetone, as their presence can lead to improve fuel properties [22, 23].

Table 1: Process Comparison

Name	Alkali-BD	BD-GLC	BD-solketal
Temperature [°C]	60	60	50
Pressure	atm	atm	atm
RSO [kmol/h]	13.16	13.16	13.16
Catalyst [kmol/h]	1.34	4.186	6.58
DMC [kmol/h]	N/A	39.48	N/A
Acetone [kmol/h]	N/A	N/A	12.75
Recycled methanol [kmol/h]	36.61	12.89	66.34
Recycled Acetone [kmol/h]	N/A	N/A	76.15
Total methanol in reactor input stream [kmol/h]	79.25	26.3	131.88
“Biodiesel” production rate, including co-products [kmol/h]	35.1	57.47	39.4

Note: “Alkali-BD” refers to the “conventional” alkali-biodiesel process, BD-GLC is the biodiesel-GLC process, BD-solketal is the biodiesel-solketal process

100,000 t/y of RSO (13.2 kmol/h) was fed to the reactor for each process. 79.2, 26.3 and 132 kmol/h of methanol were fed from methanol streams to the reactor, corresponding to 6:1, 2:1 and 10:1 (methanol: RSO) molar ratio for the Alkali-BD, BD-GLC and BD-solketal processes, respectively. The unreacted methanol was recycled in all 3 flowsheets.

Clearly, there are substantial differences between the methanol recycle rates in these three processes, which will have a significant effect on the operating costs. Notably, the recycle rate in the GLC process is approximately 1/3 that of the conventional process. This could represent a significant advantage for this process, due to the reduction in energy costs for boiling methanol. This reduction occurs because of the small amount of methanol used in the BD-GLC process (only 2:1 methanol:RSO), so only 1mol methanol per mole of RSO had to be recycled, compared with 3:1 and 9:1 molar ratios in the Alkali-BD and BD-solketal plant, respectively, which will increase process costs.

Table 2 shows the total energy consumption for each process. The BD-GLC had the lowest energy requirement among these three processes (2220 kW), followed by the conventional



alkali-BD (5400 kW). The highest energy requirement of 25800 kW was obtained for the BD-solketal process. The unit operation with the highest energy consumption in all the biodiesel processes was the distillation column as it involved methanol recovery (condenser) and glycerol separation (reboiler). It can be observed in *Table 2* that the energy required by the BD-GLC process was about 3 times lower than that of the alkali-BD process, and approximately 12 times lower than the energy requirement for the BD-solketal process. The lower energy requirement for the biodiesel-GLC process was due to the smaller methanol molar ratio used in BD-GLC process (only 2:1 methanol to RSO molar ratio), compared with 6:1 in the alkali-BD and 10:1 in BD-solketal. Note that the biodiesel-GLC process required only 3:1 DMC to RSO molar ratio, whereas the biodiesel-solketal process required 7:1 acetone to RSO molar ratio. For the same reason, the energy required by the pump to recycle the methanol was 3-fold higher in the alkali-BD process than the BD-GLC. Large amounts of the excess reactants (i.e. methanol and acetone) necessitate larger reactors, which would increase capital cost, but also increases energy requirements by increasing the heating duty of the reactors. These results clearly indicate that the overall energy consumption is largely dictated by the amount of methanol and acetone that must be recycled. Hence, the energy costs are dominated by distillation, which represents 75 - 95% of the energy cost for these processes. This observation (that the energy costs were dominated by the distillation step) has been made previously, for supercritical biodiesel processing [12].

Table 2: Energy consumption and equipment size for various process units in all processes.

<i>Item</i>	<i>Description</i>	<i>Specification</i>	<i>Alkali-BD</i>	<i>BD-GLC</i>	<i>BD-Solketal</i>	
Pumps	Recycling	Pump Duty (kW)	4.87x10 <sup>-2</sup>	1.602x10 <sup>-2</sup>	-	
Heat-Exchanger	Pre-heating	Heat Duty (kW)	250	298.8	182	
Component splitter	Neutralisation	Temperature (°C)	25	25	25	
		Pressure (kPa)	101.32	101.32	101.32	
		Heat Duty (kW)	306.3	306.9	336.4	
Reactor	Transesterification	Temperature (°C)	60	60	50	
		Pressure (kPa)	101.32	101.32	101.32	
		Heat Duty (kW)	583.7	112	649.4	
Distillation column	Methanol recovery /Glycerol separation	Temperature (°C)	28/95	28.3/64.4	56/184	64.5/81.5
		Pressure (kPa)	20/30	20/30	101.3	101.3
		Number of stages	5	5	5	15
		Reflux ratio	4	4	4	12
		Heat Duty (kW) condenser/reboiler (Total)	2082/2177 (4259)	679.5/821.6 (1501.1)	3083/ 4324 (7407)	8450/ 8761 (17211)
<b>Total energy consumption (kW)</b>			<b>5400</b>	<b>2220</b>	<b>25800</b>	

### 3.2 Economic analysis

The total capital investment (TCI) is the total direct cost, including equipment cost and project capital and the total operating cost [12, 15]. It can be observed from *Table 3* that the cost of distillation column (including condenser and reboiler) amounts to 54% of the total equipment cost for the alkali-BD process (\$0.22 M), 43% for the BD-GLC process (\$0.167 M) and 74% for the BD-solketal process (\$0.5 M).

The TCI for the BD-solketal process (\$12.87 M) was substantially higher than that of the alkali-BD (\$8.777555 M) and BD-GLC (\$7.6 M) processes. The high TCI for the biodiesel-solketal process was mainly because the reactive coupling of biodiesel and solketal requires more than one distillation column for the separation of biodiesel from other compounds to

recycle the acetone and methanol, and the distillation columns were required to handle larger amounts of methanol (10:1 methanol to RSO molar ratio) in the BD-solketal process, compared with 6:1 in alkali-BD process and 2:1 in BD-GLC.

Table 3: total cost for a biodiesel production capacity of 100,000 tonnes/y (US\$ millions).

	Alkali-BD	BD-GLC	BD-Solk
<b><i>Equipment</i></b>			
Pumps (recycling )	0.009	0.0092	-
Heat exchanger (Pre-heating & cooling)	0.0086	0.0087	0.0348
Conversional reactor	0.08	0.078	0.0974
Distillation column (main tower)	0.0751	0.05	0.13
Condenser of Distillation column	0.03	0.013	0.0472
Reboiler of Distillation column	0.0152	0.01	0.194
Total equipment cost	0.22	0.1689	0.5
water washing	0.40	0.35	0.59
FAME purification	0.5	0.44	0.74
<b><i>Project capital cost</i></b>			
Purchased Equipment	0.23	0.18	0.54
Instrumentation	0.54	0.54	0.7
Piping	0.25	0.195	0.51
Electrical	0.58	0.56	0.64
Insulation	0.052	0.056	0.102
Paint	0.02	0.013735	0.037
Equipment Setting	0.0061	0.0057	0.0155
Civil	0.044	0.04	0.093
Steel	0.023	0.019	0.05
Other	2.63	2.47	2.79
General and Administrative caused by direct labor (G and A Overheads)	0.073	0.068	0.11
Contract Fee	0.22	0.21	0.29
Contingencies	0.71	0.66	1.057
Cost of project capital cost	5.3	5.02	6.9
<b><i>Fixed capital cost</i></b>	<b>5.59</b>	<b>5.2</b>	<b>7.43</b>
Operating cost	2.25	1.65	4.1
Total capital investment (TCI)	8.75	7.6	12.87

In addition to the total investment cost, the purchasing cost of the raw material and selling price of products were included, as shown in *Table 4*. These data were also calculated based on 100,000 t/y plant capacity in *Table 5*.

Table 4: chemical prices

Item	Specification	Price (\$) /ton	Source
Biodiesel	Qualified to meet ASTM D 6751	1031	[24]
Methanol	Pure methanol 99.9%	485	[25]
Triglyceride	Rapeseed oil	813.3	[26]
Crude Glycerol	Vegetable glycerine	220	[8]
Acetone	99.9% purity	940	[8]
Dimethyl carbonate	99.9% purity	600	[27]
Glycerol carbonate	99.9% purity	15980	[28]
Solketal	99.9% purity	3000	[29]
Sodium hydroxide	-	353	[12]
DBSA catalyst	>98% purity	1000	[30]
TBD-quinidine	99.9% purity	1000	[27]
Triethyl amine	99.9% purity	1500	[27]
Acetic acid	99.9% purity	300	[27]

Table 5: revenue and profitability of each process at capacity of 100000 tonne/y (US\$ million)

Process	Alkali-BD	BD-GLC	BD-Solk
<i>direct manufacture cost</i>			
Oil	82.9	82.98	82.98
MeOH	5.4	1.78	8.92
Consumables	0.41	11.7	16.7
Acetone	0	0	0.69
DMC	0	18.68	0
(Subtotal (consumables))	(88.71)	(115.14)	(109.29)
Utility cost	0.92	0.402	2.58
Engineering and supervision	1.49	1.42	1.79
electricity	0.089	0.072	0.12
operating labour	0.65	0.34	0.35
supervision	0.3	0.3	0.31
maintenance	0.02	0.0158	0.045
operating supplies	0.162	0.0864	0.087
laboratory charges	0.105	0.056	0.057
Subtotal (direct manufacture)	3.74	2.7	5.34
<i>Indirect manufacture cost</i>			
Plant Overheads	0.335	0.33	0.35
taxes	0.224	0.224	0.224
insurance	0.097	0.097	0.097
subtotal	0.656	0.65	0.67
<b>Total production cost</b>	<b>93.2</b>	<b>118.52</b>	<b>115.4</b>
<i>Revenue</i>			
Glycerol credit	2.07	0.11	0.39
glycerol carbonate credit	0	21.6	0

solketal credit	0	0	7.23
Biodiesel revenue	103.1	103.1	103.1
<b>NPV (20y)</b>	<b>65</b>	<b>631</b>	<b>128</b>

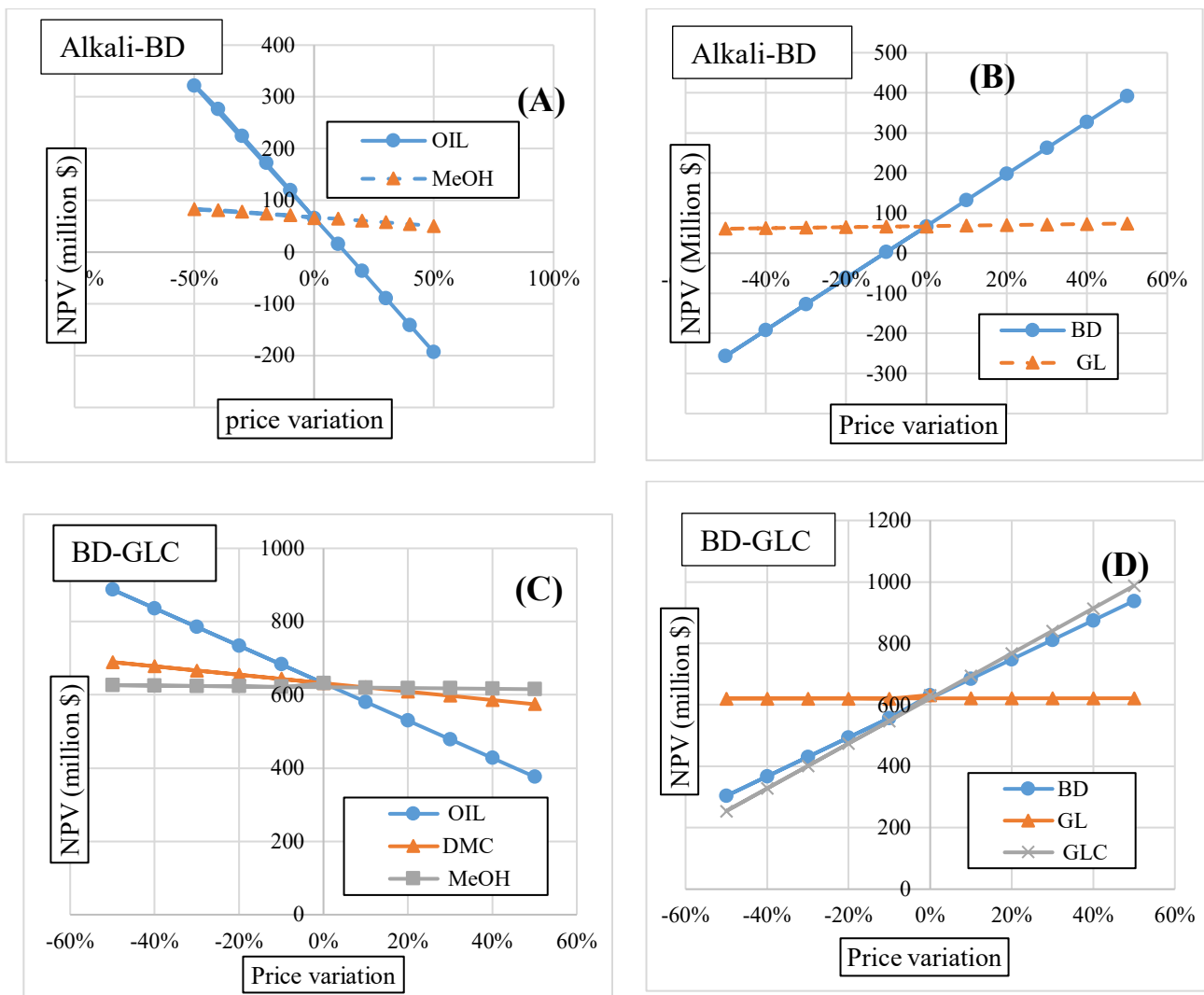
The net revenue of each process was calculated based on total production cost which includes direct and indirect costs that used in plant's building subtracted from the revenues of biodiesel credit, crude glycerol, glycerol carbonate and solketal for each process. The details of each plant building expenses and credit of outlet products are illustrated in *Table 5*.

### 3.3 Sensitivity analysis

A sensitivity analysis was performed to identify the uncertainty in net present value (NPV) for the three plants: alkali-BD, BD-GLC and BD-solketal. The current NPVs for these three biodiesel processes are \$65 M for the alkali-biodiesel process, \$128M for the biodiesel-solketal process and \$631M for the biodiesel-GLC process, as shown in *Table 5*. The purchasing prices of the raw materials (RSO, MeOH, DMC and acetone) were varied from -50% to +50% over the 20 year lifetime of the plant, with a discount rate of 15.3%. *Figure 2* shows the change in NPV as a function of variation in prices of the raw materials and the products for the biodiesel processes. The NPV of alkali-BD (*Figure 2 (a)*), BD-GLC (*Figure 2 (c)*) and BD-Solketal (*Figure 2 (e)*) were extremely sensitive to the price of RSO, decreasing by \$51 million with every 10% increase in the RSO purchase price. Change in the cost prices of methanol had little effect on the NPV: an increase in methanol purchasing price of 10% resulted only the NPV decreasing by \$3.39, \$1.1 and \$5.5 M for the alkali-BD *Figure 2 (b)*, BD-GLC *Figure 2 (d)* and BD-Solketal *Figure 2 (f)* processes, respectively. Increasing the purchasing price of dimethyl carbonate and acetone by 10% caused a reduction in the NPV by \$11.5 M (*Figure 2 (c)*) for the biodiesel-GLC process and \$0.44 M (*Figure 2 (e)*) for the biodiesel-solketal process. Changes in the RSO price had the greatest effect as it comprises the greatest part of the resultant products. It is clear from *Table 4*, that in conventional biodiesel manufacture, the cost of the main feedstock is

~80% of the product sale price. This supports the perception that the process economics of biodiesel manufacture are largely feedstock- rather than process-dependent.

Increasing the biodiesel selling price by 10% led to increases in NPV between \$63.5 and 64.9 M in the three biodiesel processes, as shown in *Figure 2(b)* for alkali-biodiesel, *Figure 2(d)* for biodiesel-GLC and *Figure 2 (f)* for biodiesel-solketal. Similarly, by increasing the selling price of the main co-products in each process (crude glycerol, GLC and solketal) by 10% the NPV increased by up to \$1.00 M for alkali-biodiesel (*Figure 2(b)*), \$73.3 million for the BD-GLC (*Figure 2(d)*), and \$ 21.4 M for the BD-Solketal process (*Figure 2 (f)*).



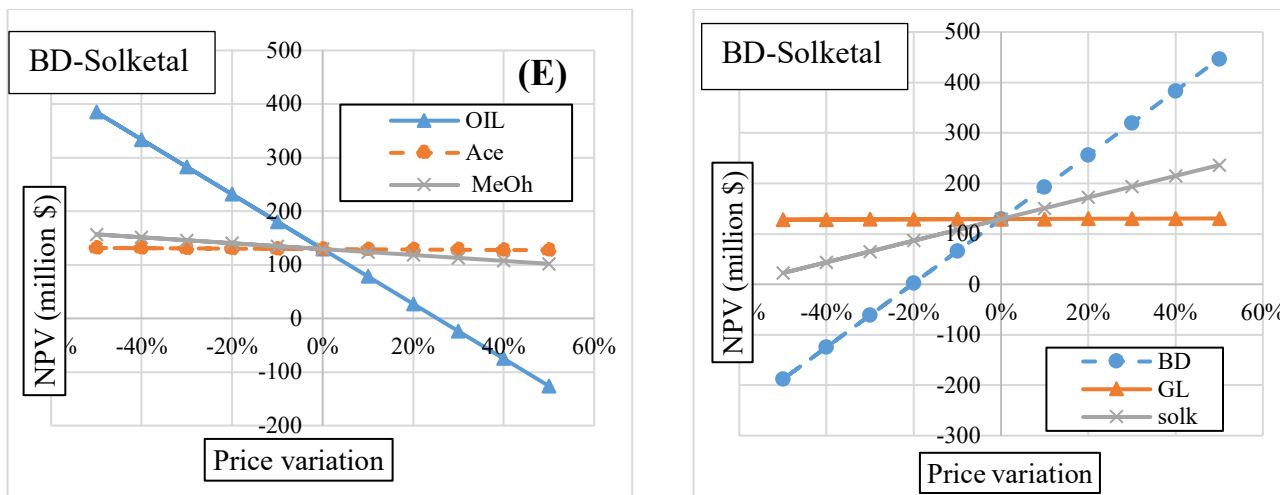


Figure 2: Net present value at different price variation for, (a) and (b) Alkali-BD plant, (c) and (d) BD-GLC plant, (e) and (f) BD-Solketal process.

Figure 2 shows that the biodiesel-GLC remains economically viable ( $NPV > 0$ ) even at over 50% increase in the biodiesel cost price (Figure 2(D)). However, the alkali-biodiesel (Figure 2(b)) and biodiesel-solketal (Figure 2(f)) processes are more sensitive to fluctuations in the price. The NPV for these processes falls below zero at biodiesel price decreases of above 10% for the alkali-biodiesel (Figure 2(b)), and biodiesel price decrease above 20% for the biodiesel-solketal process. Although the biodiesel-GLC process uses DMC, a relatively costly feedstock, the highly valuable GLC product obtained has almost 5 times the market value of solketal and commands up to a 72 times higher selling price than crude glycerol, as shown in Table 4. Consequently, this makes the BD-GLC process the most profitable among the three biodiesel processes investigated in this study. A key outcome of this analysis is that it indicates that co-production makes the processes more robust to changes in feedstock cost.

The trends in reactant and product prices lead to changes in the break-even price of the produced biodiesel, which is measured as the minimum selling price of biodiesel to achieve positive NPV. The break-even prices of biodiesel from these three process options were

\$927.9/t for the alkali-biodiesel, \$20.62/t for the biodiesel-GLC, and \$824.8/t for the biodiesel-solketal processes. These break-even prices clearly illustrate that the reactive coupling of biodiesel production with GLC formation is the most profitable process, in principle resulting in reductions of price of the produced biodiesel by a factor of 50. The next most profitable biodiesel process studied here is the reactive coupling with solketal production which has a break-even price of about \$100/t lower than the conventional alkali-biodiesel process.

### ***3.4 Capital and annual production costs for different plant capacities***

As the BD-GLC process was significantly more economically viable, it was investigated further. There are no such plants operating currently, and it is unlikely that the first plant constructed would operate at 100 kt/y scale due to the perceived risk in any new process. Hence, the effects of production capacity on profitability were investigated for this process. The three capacities studied were: 10,000, 40,000 and 100,000 t/y. The total capital investment were \$5.3M for the 10,000 t/y, \$5.7 M for the 40,000 t/y and \$7.65 M for the 100,000 t/y, as shown in *Table 6*. The largest plant capacity was significantly more profitable than those of the small and medium plants, the NPV and the net revenue of the largest capacity plant was greater than that of the small capacity plant by a factor of 18, and more than the medium capacity plant by a factor of 2.7. The net revenue from large capacity plant is more than medium plant by 2.6 fold and more than 15 time of small plant.

Table 6: effect of production capacity on BD-GLC production cost

Capacity (t/y)	10,000	40,000	100,000
TCI (\$ M)	5.3	5.7	7.65
NPV (20y)	34.6	228.5	631
Net revenue (\$M)	6.5	38.05	100.28
Operating cost (\$M)	1.35	1.47	1.67



## **4 Conclusions**

Techno-economic analysis was performed for biodiesel production via three processes: the conventional alkali-catalysed , reactive coupling with co-production of glycerol carbonate (GLC) and reactive coupling with co-production of solketal.

Aspen HYSYS was used to simulate these processes, based on a production capacity of 100,000 t/y and conversion of triglyceride to FAME of 98%. Economic evaluations showed that the capital investment cost, energy consumption and NPVs for the three biodiesel processes were as (i) Conventional biodiesel process: \$7.84 M capital investment, 5.39 MW per annum and \$65 M (NPV), (ii) Reactive coupling with GLC formation: \$6.84 M capital investment, 2.2MW per annum and \$631 M (NPV) and (iii) Reactive coupling with solketal production: \$11.53 M, 25.8MW for the process and \$128M (NPV).

As the GLC co-production process was significantly more profitable than the other processes, the effect of scale was investigated for this process alone. The NPV increased approximately linearly with scale. Operating the GLC process at 100,000 kt/y (NPV \$631 M) was found to be significantly more profitable than the lower production capacities: 40,000 t/y (\$228.5 M) and 10,000 t/y (\$34.6 M). This was because the TCIs changed little with increasing production capacity.

Overall, it is clear that co-production of solketal and GLC is economically viable, and an improvement on production of biodiesel as a sole product. Beyond this, the co-production of GLC is substantially more profitable than solketal, with an NPV 4 times higher.

## **Acknowledgement**

The authors would like to thank the Higher Committee for Education Development in Iraq (HCED) for their financial support.

## References

1. Combs, S., *The Energy Report in Biodiesel*. 2008, Texas Comptroller
2. Demirbas, A., *A Realistic Fuel Alternative for Diesel Engines*. 1 ed. 2008: Springer London. 1-208.
3. Helwani, Z., et al., *Technologies for production of biodiesel focusing on green catalytic techniques: A review*. Fuel Processing Technology 2009. **90** p. 1502–1514.
4. Knothe, G., J. Krahl, and J. Van Gerpen, *The Biodiesel Handbook: Second Edition*. The Biodiesel Handbook: Second Edition. 2010. 1-501.
5. Kumar, D., et al., *Cement wastes as transesterification catalysts for the production of biodiesel from Karanja oil*. Journal of Cleaner Production, 2018. **183**: p. 26-34.
6. Ayoub, M. and A.Z. Abdullah, *Critical review on the current scenario and significance of crude glycerol resulting from biodiesel industry towards more sustainable renewable energy industry*. Renewable and Sustainable Energy Reviews, 2012. **16**(5): p. 2671-2686.
7. Quispe, C.A.G., C.J.R. Coronado, and J.A. Carvalho Jr, *Glycerol: Production, consumption, prices, characterization and new trends in combustion*. Renewable and Sustainable Energy Reviews, 2013. **27**: p. 475-493.
8. ICIS, *Indicative Chemical Prices A-Z*, in *ICIS Chemical Business magazine* 2018, ICIS Resource centre: New York.
9. Eze, V.C. and A.P. Harvey, *Continuous reactive coupling of glycerol and acetone – A strategy for triglyceride transesterification and in-situ valorisation of glycerol by-product*. Chemical Engineering Journal, 2018. **347**: p. 41-51.
10. Lee, Y., et al., *Efficient simultaneous production of biodiesel and glycerol carbonate via statistical optimization*. Journal of Industrial and Engineering Chemistry, 2017. **51**: p. 49-53.
11. Dhawan, M.S. and G.D. Yadav, *Insight into a catalytic process for simultaneous production of biodiesel and glycerol carbonate from triglycerides*. Catalysis Today, 2017.
12. Sakdasri, W., R. Sawangkeaw, and S. Ngamprasertsith, *Techno-economic analysis of biodiesel production from palm oil with supercritical methanol at a low molar ratio*. Energy, 2018. **152**: p. 144-153.
13. Yusuf, N.N.A.N. and S.K. Kamarudin, *Techno-economic analysis of biodiesel production from *Jatropha curcas* via a supercritical methanol process*. Energy Conversion and Management, 2013. **75**: p. 710-717.
14. Vlysidis, A., et al., *A techno-economic analysis of biodiesel biorefineries: Assessment of integrated designs for the co-production of fuels and chemicals*. Energy, 2011. **36**(8): p. 4671-4683.

15. Lee, S., D. Posarac, and N. Ellis, *Process simulation and economic analysis of biodiesel production processes using fresh and waste vegetable oil and supercritical methanol*. Chemical Engineering Research and Design, 2011. **89**(12): p. 2626-2642.
16. Apostolakou, A.A., et al., *Techno-economic analysis of a biodiesel production process from vegetable oils*. Fuel Processing Technology, 2009. **90**(7): p. 1023-1031.
17. Zhang, Y., et al., *Biodiesel production from waste cooking oil: 1. Process design and technological assessment*. Bioresource Technology, 2003. **89**(1): p. 1-16.
18. Freedman, B., E.H. Pryde, and T.L. Mounts, *Variables affecting the yields of fatty esters from transesterified vegetable oils*. Journal of the American Oil Chemists Society, 1984. **61**(10): p. 1638-1643.
19. Al-Saadi, L.S., V.C. Eze, and A.P. Harvey, *A reactive coupling process for co-production of solketal and biodiesel*. Green Processing and Synthesis, 2019. **8**(1): p. 516-524.
20. Al-Saadi, L.S., V.C. Eze, and A.P. Harvey, *Experimental Determination of Optimal Conditions for Reactive Coupling of Biodiesel Production With in situ Glycerol Carbonate Formation in a Triglyceride Transesterification Process*. Frontiers in Chemistry, 2018. **6**(625).
21. West, A.H., D. Posarac, and N. Ellis, *Assessment of four biodiesel production processes using HYSYS.Plant*. Bioresource Technology, 2008. **99**(14): p. 6587-6601.
22. Rajesh Kumar, B. and S. Saravanan, *Partially premixed low temperature combustion using dimethyl carbonate (DMC) in a DI diesel engine for favorable smoke/NOx emissions*. Fuel, 2016. **180**: p. 396-406.
23. Elfasakhany, A., *Performance and emissions analysis on using acetone-gasoline fuel blends in spark-ignition engine*. Engineering Science and Technology, an International Journal, 2016. **19**(3): p. 1224-1232.
24. CARD, *Historical Biodiesel Operating Margins*. 2018.
25. Methanex, *Methanex posts regional contract methanol prices for North America, Europe and Asia*. . 2018.
26. International Monetary Fund, *commodity price of rapeseed oil 2018*, Index mundi: united state.
27. Alibaba, *chemicals categories*. 2018, [www.alibaba.com](http://www.alibaba.com).
28. indiamart, *Industrial Chemicals & Supplies*. 2018.
29. Qyresearch, *Chemical raw materials*. 2018, A US Professional Market Research Publisher
30. made-in-china.com, *chemical categories*. 2018, third-party B2B e-commerce platform in China China.

