Techno-economic analysis of a biodiesel production process from vegetable oils

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A B S T R A C T

Biodiesel, which is defined as the monoalkyl esters of long chain fatty acids derived from a renewable lipid feedstock, has received considerable attention worldwide as a medium-term alternative to diesel fuel obtained from petroleum. Biodiesel can be produced by the transesterification of vegetable oils or animal fats using short-chain alcohols in the presence of a suitable catalyst and glycerol is the only byproduct obtained in significant quantities. In this work a techno-economic analysis of a process that produces biodiesel from vegetable oils is presented with the aim to investigate the dependence of the critical profitability indicators on the production capacity.

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1. Introduction

Biodiesel, which is defined as the monoalkyl esters of long chain fatty acids derived from a renewable lipid feedstock, has received considerable attention worldwide as a medium-term alternative to diesel fuel obtained from petroleum. Biodiesel can be produced by the transesterification of oils with short-chain alcohols, such as methanol, or by the esterification of free fatty acids (FFA). The transesterification reaction is taking place in the presence of a suitable catalyst such as alkali or acid and glycerol is the only byproduct produced. Approximately 100 kg of oil reacts with 10 kg of methanol to produce 100 kg of biodiesel and 10 kg of glycerol. Biodiesel can also be produced at supercritical conditions in the absence of any catalyst or through biochemical routes based on the use of enzymes [1].

The main advantages of biodiesel are a) it is made from renewable resources that can be produced domestically b) it produces less carbon monoxide, particulates, and sulfur dioxide emissions c) it produces 78% less carbon dioxide and d) it is biodegradable, non-toxic and safer to handle. The main disadvantages are its high price, the increased nitrogen oxide emissions and the concerns about its impact on the durability of diesel engines when used in pure form (B100).

As part of a range of measures to reduce greenhouse gas emissions, the EU is encouraging the use of biofuels (both biodiesel and bioethanol) and the EU Biofuels Directive requires 5.75% of the energy for transport to come from renewable sources by the end of 2010 (10% by 2020). Transport fuels account for around 25% of EUs greenhouse gas emissions and demand for diesel and petrol is fast rising and, by 2020, fuel consumption is estimated to reach 325 Mt. Europe and US

biodiesel producers, in an effort to reduce production costs by taking advantage of the economies of scale, transition to larger biodiesel production plants [2]. While currently the average biodiesel production plant capacity is around 40 kt/year it is envisaged that plants with capacities as large as 800 kt/year will be built in the near future [3].

Europe has dominated the biodiesel industry up-to-date with 90% of global production where tax exemptions and national targets introduced through EU directives are driving the market. The EU produced 2.4 Mt of biofuels in 2004, amounting to 0.8% of EU petrol and diesel consumption. Germany has the highest consumption of biodiesel at 1.1 Mt in 2004, 2 Mt in 2005 and 3.2 Mt in 2006. Rapeseed oil is the main biodiesel feedstock, constituting just over 20% of EU25 total oilseed production.

A number of studies have been published that investigate the economics of biodiesel production processes and a review of the more recent and relevant works is presented in the sequel. The review is not intended to be comprehensive and the interested reader is referred to the reviewed papers for references and critical reviews of other relevant works.

Zhang et al. [4,5] present an economic study of a biodiesel production plant with a capacity of 8 kt/year using waste cooking oils as a raw material. They compare four different available production technologies and conclude that the alkali-catalyzed, transesterification features significantly lower total capital investment (TCI). The process simulation of alkali-catalyzed transesterification process is based on the early experiments by Freedman et al. [6,7]. They further conclude that based on after-tax rate of return and break-even price of biodiesel, the acid-catalyzed processes were economically competitive alternatives to the alkali process for biodiesel production. Sensitivity analysis of different processes for biodiesel production shows that plant capacity, the price of waste cooking oil and the price of biodiesel were the major factors affecting the economic feasibility

Nomenclature

A	heat transfer area (m^2)
BM	bare module
C^0	fixed on board equipment cost
c	unit cost
D_C	distillation column diameter (m)
DCFR	discounted cash flow rate of return
EU	European Union
FCI	fixed capital investment
FFA	free fatty acids
fob	fixed on board
H_C	distillation column height (m)
kt	thousand metric tons
Mt	million metric tons
MW	molecular weight
N	number of trays in a distillation column
P	pressure (kPa or bar)
Q	volumetric flowrate (m^3/h)
R	recovery (%)
ROI	return on investment
T	temperature (K)
TCI	total capital investment
τ_R	residence time (h)
V	volume (m^3) or vapor mass flowrate (kg/s)
X	conversion

of the biodiesel production. Zhang et al. use the commercially available simulator HYSYS for solving the pertinent material and energy balances.

A comprehensive study is presented by Haas et al. [8] where the economics of a 33.1 kt/year plant, that produces biodiesel from crude degummed soy oil using alkali-catalyzed transesterification technology, is presented. They conclude that the total capital cost is of the order of M$11.35 while soy oil cost accounts for 88% of the unit production cost. All steady state simulations are performed in the commercially available simulator ASPEN PLUS. The process simulation of alkali-catalyzed transesterification process is also based on the early experiments by Freedman et al. [6,7]. Harding et al. [9] also use ASPEN PLUS to simulate production processes that use either inorganic or biological catalysis for the production of biodiesel by transesterification using rapeseed oil. They present interesting results that are related to their environmental impact. They conclude that the enzymatic production route is environmentally more favorable as expected.

Marchetti et al. [10] compare the economics of alkali-catalyzed, acid-catalyzed and solid resin-catalyzed process for producing approximately 36 kt/year of biodiesel from spent oils with 5% free fatty acids. They use the commercial simulator SuperPro Designer in order to simulate the relative processes. They conclude that in all cases raw materials cost is the most important contribution to the unit production cost (which is estimated around $0.5–0.52/kg). The alkali-catalyzed based process appears to be the most economically attractive alternative followed by the heterogeneous catalyst based process.
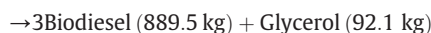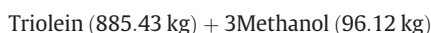
Marchetti and Errazu [11] compare the economics of a biodiesel plant that operates at supercritical conditions with the economics of biodiesel plants that use classical catalytic production routes. They conclude that all profitability indicators are worse for the plant that operates at supercritical conditions due to higher operating costs involved in the supercritical plant which can partially be attributed to the high pressure (70 bar) and high temperature (200 °C) of operation of the reactor(s).

West et al. [12] compare the economics of four different technologies for producing biodiesel: homogeneous alkali catalyst processes, homogeneous basic catalyst processes, heterogeneous catalyst processes and supercritical processes. For a production capacity of 8 kt/year they conclude that the heterogeneous catalytic process is the most economically attractive process and at the same time the least complex. All other processes appear not to be profitable.
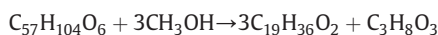
All previous studies investigate either the feasibility of an existing technology [4,5,8] or compare alternative technologies [10–12] for a predefined plant capacity. The aim of this work is to investigate the effect of production capacity on biodiesel plant profitability. The case of biodiesel production from oils that are free from FFA and employ homogeneous alkali catalysts will be considered.

2. Representative process description

The process used in this study is shown in Fig. 1. The selected process follows closely with the one used previously by Haas et al. [8] while many technical details can be found in other literature sources [13–16] as well. The oil (triolein) reacts with excess methanol at the presence of homogeneous alkali catalyst (sodium methoxide) according to the reaction:

Triolein (885.43 kg) + 3Methanol (96.12 kg)

→3Biodiesel (889.5 kg) + Glycerol (92.1 kg)

or

$$C_{57}H_{104}O_6 + 3CH_3OH \rightarrow 3C_{19}H_{36}O_2 + C_3H_8O_3$$

Triolein is used to represent rapeseed oil. 100% excess methanol (molar ratio 6:1) is used.

Rapeseed oil (stream 1) together with excess methanol and catalyst (stream 2 – sodium methoxide – 1% of the oil weight) are fed to the first of two well-mixed reactors (R-101). It is assumed that the conversion of the rapeseed oil to biodiesel at atmospheric pressure and 60 °C is 90% if the residence time is selected to be 1 h (both are based on the work published by Freedman et al. [6,7], see also discussion in literature [13–16]). The reactor effluent (stream 4) is fed to a centrifugal separator (FF-101) where the unreacted oil and the biodiesel produced (stream 5) are separated from glycerol (stream 6).

Stream 5 is fed to the second well-mixed reactor (R-102) together with the remaining methanol and catalyst (20% of the overall amount of methanol and catalyst) and at atmospheric pressure, 60 °C and a residence of 1 h 90% conversion of the remaining triolein is achieved resulting in an almost complete combined conversion of triolein (99%). The reactor effluent (stream 7) is fed to a centrifugal separator (FF-102) where the unreacted oil and the biodiesel produced (stream 9) are separated from glycerol (stream 8).

Stream 9 is fed to a mixing tank V-101 together with process (wash) water (stream 10) and HCL solution (stream 3) so as to neutralize the catalyst and convert any soap to FFA. The wash and pH adjustment tank effluent is fed to a centrifuge (FF-103) where biodiesel is recovered together with small amounts of water (stream 12). The final purification of the biodiesel is achieved in a flash drum (D-101) that operates under vacuum (5 kPa).

Streams 6, 8, 13 and 15 that contain significant amounts of methanol, glycerol, and water as well as FFA, soaps and salts are fed to a mixing tank (V-102). The combined stream (stream 16) is first treated with HCL (stream 17) to convert any soap to FFA and then (stream 18) is fed to a centrifuge to remove the FFA (stream 20). The glycerol and methanol rich stream (stream 19) is first neutralized with NaOH (stream 21) in the pH adjustment tank V-103 and the resulting stream (stream 22) is fed to a distillation column (T-101) which

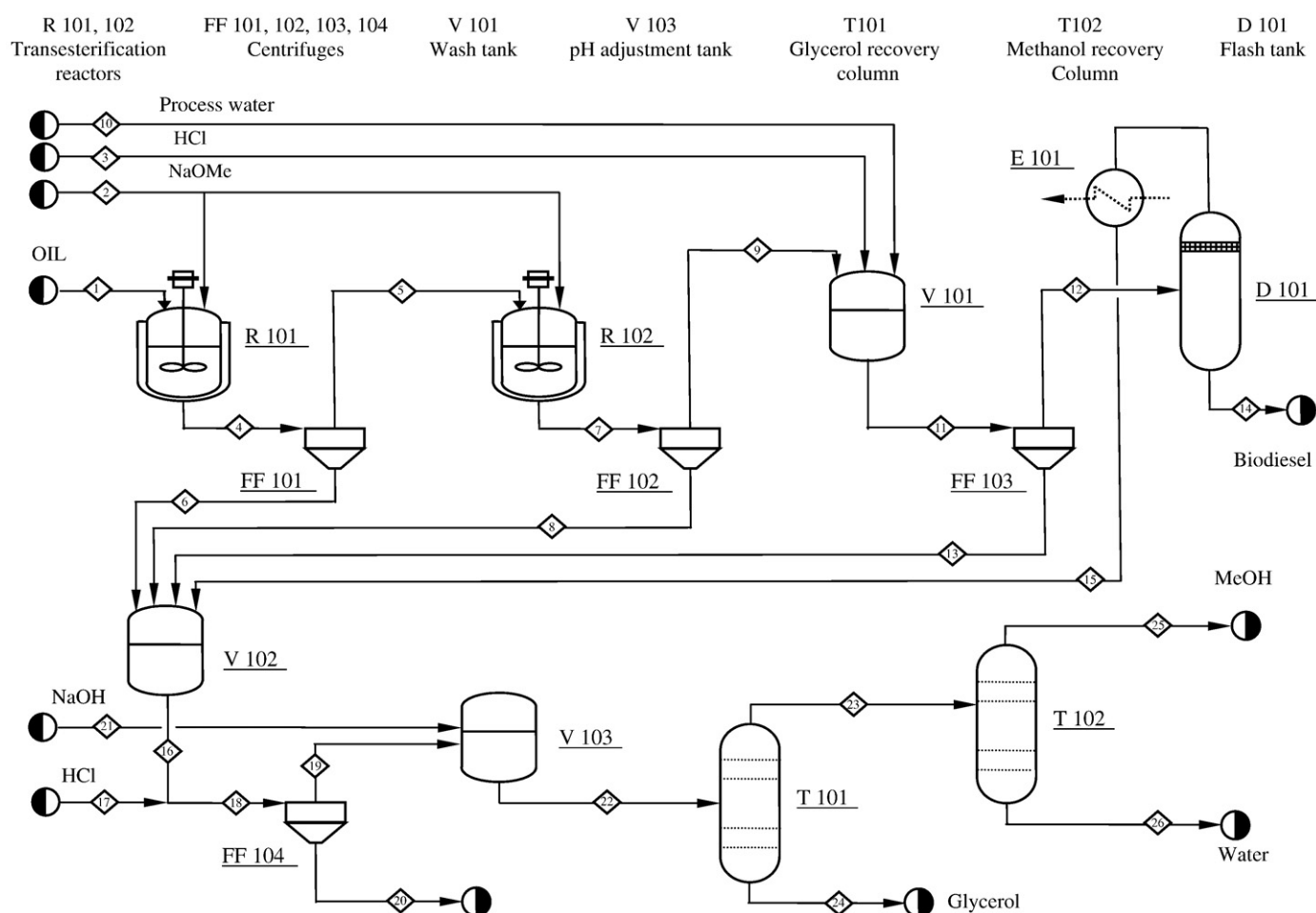| R 101, 102	FF 101, 102, 103, 104	V 101	V 103	T101	T102	D 101
Transesterification reactors	Centrifuges	Wash tank	pH adjustment tank	Glycerol recovery column	Methanol recovery Column	Flash tank

Fig. 1. Process flow diagram of the biodiesel production plant.

operates slightly above atmospheric pressure. All glycerol is removed as the bottom product of T-101 (stream 24) which is 80% w/w in glycerol and the remaining is predominantly water. The top product (stream 23) consists of water and methanol and is fed to a second distillation column which operates at 50 kPa. Almost pure (99.9% mol) methanol is obtained as top product (stream 25) which is recycled back to the methanol storage tank. The bottom product (stream 26) is also pure water that can be fed to the wash tanks.

3. Process steady state simulation

The commercial simulator HYSYS [17] is used to solve the steady state material and energy balances. In order to achieve that the HYSYS thermophysical properties database needs to be extended to incorporate triolein that is not available in the HYSYS component library.

Data for triolein are very scarce in the literature as triolein decomposes at temperatures far below its boiling point. HYSYS requires the knowledge of the molecular weight, normal boiling point, liquid density, critical temperature, critical pressure, critical volume and the acentric factor. In the cases where not all of them are available HYSYS can estimate the remaining using classical group contribution methods.

The normal boiling point has been estimated by Tang et al. [18] using the method of Dohrn and Brunner which requires only the liquid molar volume and vapor pressure data. They estimated the boiling temperature as $T_{nb} = 879.9$ K (606.8 °C). It is interesting to note that many previous studies have used the boiling point of triolein that is available in an ASPEN Plus commercial simulator which is given as $T_{nb} = 1119.15$ K (846 °C). This value appears to be an overestimation of the unknown normal boiling point. Vapor pressure data for triolein

can be found in [19] and [20] from which the following equation for the calculation of vapor pressure of triolein can be obtained:

$$\ln P^{vp}[\text{bar}] = 6.596 - \frac{5407.3.3}{T[\text{K}]} \tag{1}$$

From Eq. (1) we can obtain an estimate of the normal boiling point temperature $T_{nb} = 821.5$ K (548.3 °C) which supports our argument that the value available in ASPEN Plus and used by many authors is possibly an overestimation of the actual value.

Many authors have used Group Contribution methods to estimate the critical pressure and critical temperature of triolein. Tang et al. [18] have estimated the critical temperature $T_C = 954.1$ K and the critical pressure $P_C = 0.36$ MPa. These estimations are in agreement with the estimations of Weber et al. [20] for the critical temperature $T_C = 947.1$ K and the critical pressure $P_C = 0.468$ MPa. Vázquez et al. [21] estimated a slightly larger value for the critical temperature $T_C = 1020$ K and for the critical pressure $P_C = 0.45$ MPa. It is also interesting to note that the critical temperature estimation for triolein that is available in ASPEN Plus database is $T_C = 1640$ K which is again considered to be an overestimation. The critical pressure estimation for triolein that is available in ASPEN Plus database is $P_C = 0.47$ MPa which is close to the estimations by Weber et al. and Vázquez et al. Tang et al. [18] also report values for the acentric factor $\omega = 1.6862$ which agrees exactly with the value given by Weber et al. [20].

In this work the values for the normal boiling point temperature, critical temperature, critical pressure, liquid molar volume (0.9717 m³ kmol⁻¹) and acentric factor given by Tang et al. [18] are used in order to define triolein in the HYSYS component database. The

NRTL activity model is used together with the UNIFAC VLE-LLE method for estimating all missing coefficients. The methanol–water and glycerol–water VLE models were tested using equilibrium data available in Perry's Handbook [22]. On overall the thermodynamic model accuracy is considered acceptable for the purposes of a preliminary cost estimation study.

The HYSYS model developed is shown in Fig. 2. Transesterification reactors are modeled as Conversion Reactor, centrifuges as Component Splitter, mixing tanks as Tank, flash drums as Separator and distillation columns as Distillation Column. The conversion to the reactors is set to 90% conversion of triolein following the experimental results by published by Freedman et al. [6,7], (see also discussion in literature [13–16]). Centrifuges are assumed to achieve 99% recovery of components. In FF-101 and FF-102 methanol is assumed to distribute by 60% in the biodiesel phase and 40% in the glycerol phase. In FF-103 methanol is assumed to distribute by 10% in the biodiesel phase and in FF-104 by 100% in the glycerol phase. The distillation columns are designed using McCabe Thiele diagram for the second column (T-102) and stage-by-stage calculations for the first column (T-101).

4. Equipment specifications, design and purchase cost estimation

The process design is based on specifications obtained from the literature [8,13–16].

4.1. Transesterification reactors

The design of the transesterification reactors is based on the assumption that using 100% excess methanol and 1% catalyst (based on the weight of the oil) and a residence time (τ_R) of 1 h a conversion of 90% of oil can be achieved at 60 °C and low pressure. The volumetric flowrate (Q) of the incoming streams is calculated using HYSYS and the reactor volumes (V) are calculated using the following equation:

$$V = \frac{\tau_R Q}{0.8} \tag{2}$$

The fob cost (C_R^0) of the jacketed, agitated, well-mixed reactors is calculated based on the following equation [23]:

$$C_R^0 = 15000 \, V^{0.55} \tag{3}$$

where V is the reactor volume in m^3. Eq. (3) applies for construction from stainless steel and reactor volumes between 0.1 m^3 and 20 m^3. All cost figures will be in US$ at 2007.

4.2. Centrifuges

Centrifuge fob (C_{FF}^0) cost can be calculated using the following equation:

$$C_{FF}^0 = 28100 \, Q^{0.574} \tag{4}$$

where Q is the incoming stream flowrate in m^3/h. Eq. (4) applies for construction from stainless steel and incoming flowrates between 0.1 and 100 m^3/h. The equation is valid for separating liquid mixtures that appear in biodiesel production and involve biodiesel, oil, glycerol, water and methanol as it has been derived from vendor quotations for the specific separations. The incoming stream flowrate is again obtained from HYSYS.

4.3. Mixing and pH adjustment tanks

The fob cost (C_V^0) of mixing and pH adjustment tanks can be calculated using the following equation:

$$C_V^0 = 12080 \, V^{0.525} \tag{5}$$

where V is the tank volume in m^3 calculated from Eq. (2). Eq. (5) applies for construction from stainless steel and tank volumes between 0.1 m^3 and 30 m^3. All tanks are designed using Eq. (2) and a residence time of 1 h.

4.4. Flash drums

The following equation is used to calculate the fob cost of flash drums (C_D^0) which is derived from the classical Guthrie's correlations [23]

$$C_D^0 = 6500 \, V^{0.62} \tag{6}$$

where V is the tank volume in m^3. Eq. (6) applies for construction from stainless steel and tank volumes between 0.1 m^3 and 70 m^3.

4.5. Heat exchangers

The following equation is used to calculate the fob cost of shell and tube heat exchangers (C_E^0) which is derived from the classical Guthrie's correlations [23]:

$$C_E^0 = 2320 \, A^{0.65} \tag{7}$$

where A is the heat transfer area in m^2. Eq. (7) applies for construction from carbon steel and heat transfer areas between 10 m^2 and 1000 m^2. The installed equipment cost for the case of heat exchangers fabricated from stainless steel is calculated as follows:

$$C_E = (2.29 + 1.78 f_d) \, C_E^0 \tag{8}$$

where f_d is the design correction coefficient which obtains the value of 1.35 for kettle reboilers and the value of 1 for floating head heat exchangers. In order to calculate the heat transfer area an overall heat transfer coefficient of 0.5 $kW·m^{-2} K^{-1}$ is assumed for all heat exchangers apart from reboilers for which an overall heat transfer coefficient of 1 $kW·m^{-2}K^{-1}$ is assumed. All stream temperatures are obtained from HYSYS.

4.6. Distillation columns

The fob cost of distillation columns (C_T^0) is obtained from the classical Guthrie's correlation:

$$C_T^0 = 4555 \, H_C^{0.81} \, D_C^{1.05} \tag{9}$$

where D_C is the column diameter in m and H_C is the column height in m. The column height is calculated as follows:

$$H_C = 1.2 \, TS(N - 1) \tag{10}$$

where TS is the tray spacing in m and N the number of real trays. The column diameter is calculated as:

$$D_C = \sqrt{\frac{4A_n}{0.88\pi}} \tag{11}$$

and the net area A_n as:

$$A_n = \max_{\text{all trays}} \left[\frac{V}{2\sqrt{\rho_V}} \right] \tag{12}$$

The values of the vapor mass flowrate (V) and vapor mass density (ρ_V) are obtained from HYSYS at each tray of the distillation column.

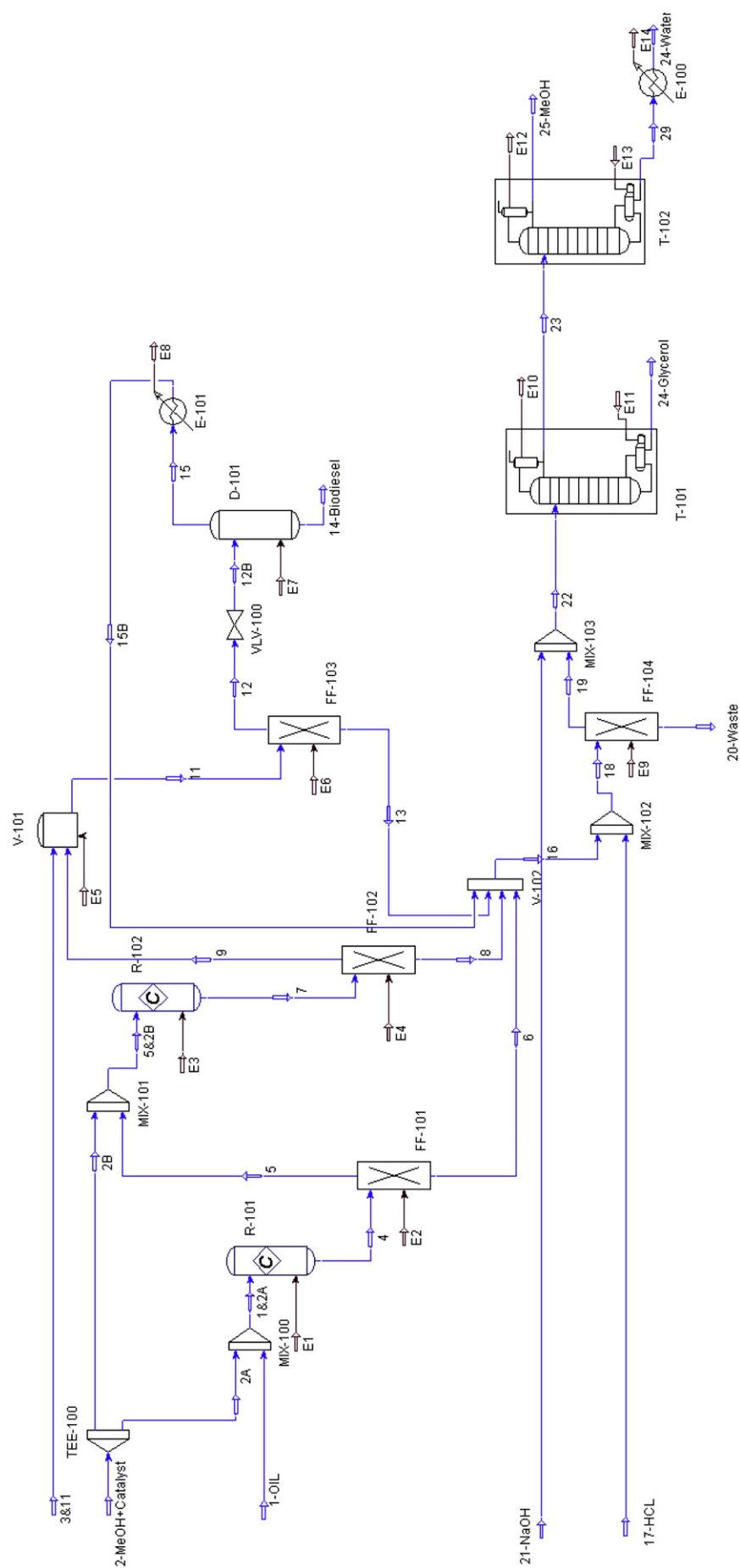Fig. 2. HYSYS model of the biodiesel production plant.

Table 1
Summary of equipment specifications and installed equipment cost calculation (plant capacity: 50 kt/year).

Equipment	Specification	Char. size	f.o.b. cost (M$)	BM factor	Installed eq. cost (M$)
Reactors					
R-101	1 h residence time	$V=9.1$ m^3	0.101	2.8	0.283
R-102	1 h residence time	$V=8.8$ m^3	0.099		0.278
Centrifuges					
FF-101		$Q=7.2$ m^3/h	0.068	1.3	0.088
FF-102		$Q=7$ m^3/h	0.067		0.087
FF-103		$Q=7.8$ m^3/h	0.071		0.092
FF-104		$Q=2.8$ m^3/h	0.040		0.052
Mixing and pH adjustment tanks					
V-101	1 h residence time	$V=9.7$ m^3	0.040	2.8	0.111
V-102	1 h residence time	$V=3.5$ m^3	0.024		0.066
V-103	1 h residence time	$V=3.3$ m^3	0.023		0.063
Flash drum					
D-101	0.2 h liquid product residence time	$V=3.3$ m^3	0.015	2.8	0.043
Distillation columns					
T-101	$N=23$, $E_O=0.7$	$D_C=0.6$ m $H_C=10.8$ m			0.184
T-102	$N=29$, $E_O=0.7$	$D_C=0.84$ m $H_C=13.7$ m			0.105
Heat exchangers					
E-101	$U=0.5$ kW/(m^2 K)	$A=27$ m^2	0.014	2.2	0.031
Reboilers					
T-101	$U=1$ kW/(m^2 K)	$A=37.7$ m^2	0.024	4.7	0.113
T-102	$U=1$ kW/(m^2 K)	$A=12$ m^2	0.012		0.053
Condensers					
T-101	$U=0.5$ kW/(m^2 K)	$A=34$ m^2	0.023	4.07	0.092
T-102	$U=0.5$ kW/(m^2 K)	$A=106$ m^2	0.048		0.195
Storage tanks					
Rapeseed oil	4 weeks storage capacity	$V=5515$ m^3	0.770	1.2	0.923
Methanol		$V=1350$ m^3	0.280		0.335
Biodiesel		$V=5635$ m^3	0.781		0.937
Glycerol		$V=676$ m^3	0.172		0.207
Total installed equipment cost (1.2 × sum of installed eq. cost)					5.2
Fixed capital investment (FCI ≈ 2 × total installed eq. cost)					M$9.4

The cost of the installed sieve trays is obtained from the equation:

$$C_{TR} = 380 N^{0.81} D_C^{1.55} \qquad (13)$$

4.7. Storage tanks

The fob cost of field erected, stainless steel storage tanks can be calculated from the equation [23]:

$$C_{ST}^0 = 250000 + 94.2V \qquad (14)$$

where V is the storage tank volume in m^3. Eq. (14) applies for construction from stainless steel and tank volumes between 2×10^3 m^3 and 50×10^3 m^3. For smaller tanks the following equation should be used:

$$C_{ST}^0 = 65000 + 158.7V \qquad (15)$$

The required storage capacity is obtained by assuming that it is required to have available storage facilities to support production and product storage for four weeks.

Using Eqs. (1)–(15) the fixed capital investment (FCI) of a biodiesel plant, producing 50 kt/year biodiesel from rapeseed oil (free of FFA), is calculated to be of the order of M$9.4. Table 1 summarizes the calculations performed. The plant is considered to be a new (grass roots) plant and as a result the FCI is calculated to be 1.8 times the overall cost of the installed equipment [23]. Storage tanks, which account for 46% of the FCI, and transesterification reactors, which account for 11% of the FCI, are the main contributors to the FCI.

5. Estimation of the operating costs

5.1. Raw materials cost

The raw materials cost (C_{RM}) consists of the cost of rapeseed oil, the cost of methanol, the cost of catalyst and the cost of HCL and NaOH solutions.

The estimation of the cost of the rapeseed oil is based on the overall material balance of the process. The rapeseed oil required to produce 50 kt/year of biodiesel is 50.8 kt/year as the overall conversion is 99% and the biodiesel recovery 99%. The cost of rapeseed oil at the EU market was $1000/t at the beginning of the year 2007 and was approaching $1150/t at the end of the same year. For the purposes of this study a mean value of $1100/t is assumed. As a result the cost of the rapeseed oil is M$55.866.

The methanol required is 5.4 kt/year and the price of methanol is $300/t. As a result the cost of methanol is M$1.62. Using similar arguments we can calculate the cost of the catalyst (M$0.313) and the cost of the HCL and NaOH solutions (M$0.1). The overall raw material cost is then $C_{RM}=$ M$57.9. It is interesting to note that the cost of oil accounts for 96.5% of the raw materials cost.

5.2. Utilities and operating labor costs

The cost of the utilities (C_{UT}) consists of the cost of electricity and the cost of steam. It is estimated through the energy balances of the process that the utilities consumption are approximately 30 kWh/t biodiesel produced and 1.4 GJ/t biodiesel produced of electricity and of heating utility, respectively. Assuming a price of $0.15/kWh for electricity and $10/GJ of heating utility we can calculate that for the production of 50 kt/year of biodiesel $C_{UT}=$ M$0.91.

The operating labor cost (C_{OL}) is based on the estimation that a plant producing 50 kt/year biodiesel will require 15 operators and an average cost of $40,000/year per operator. As a result the operating labor cost is $C_{OL}=$ M$0.6/year.

5.3. Unit production cost calculation

Table 2 summarizes the calculation of the total production cost and the unit production cost [24]. The total production cost for a biodiesel plant that produces 50 kt/year of biodiesel is estimated to be M$65.9 while the unit production cost is estimated to be $1.15/L ($4.353/gal). This is approximately €0.767/L which is slightly lower than the

Table 2
Summary of operating cost calculation (plant capacity: 50 kt/year).

|	Calculation	M$	%
Cost item			
1. Raw materials cost (C_{RM})	From material balances	57.900	87.86
2. Miscellaneous materials	1% of FCI	0.094	0.14
3. Utilities cost (C_{UT})	from material balances	0.910	1.38
Variable costs	(1) + (2) + (3)	58.904	
4. Maintenance	10% of FCI	0.940	1.43
5. Operating labor (C_{OL})	manning estimates	0.600	0.91
6. Lab costs	20% of (5)	0.120	0.18
7. Supervision	20% of (5)	0.120	0.18
8. Overheads	50% of (5)	0.300	0.45
9. Capital charges	15% of FCI	1.406	2.13
10. Insurance, local taxes and royalties	4% of FCI	0.375	0.57
Fixed costs	(4) + (5) + ... + (10)	3.861	
Direct production costs	Variable costs + Fixed costs	62.765	
11. General overheads + R&D	5% of the direct prod. costs	3.140	4.80
Annual production costs	(1) + (2) + ... + (11)	65.9	100.00
Unit production cost	$65.9/50 kg biodiesel = $1.150/L biodiesel		
	= $1.318/kg biodiesel		
	≈ €0.766/ L biodiesel		
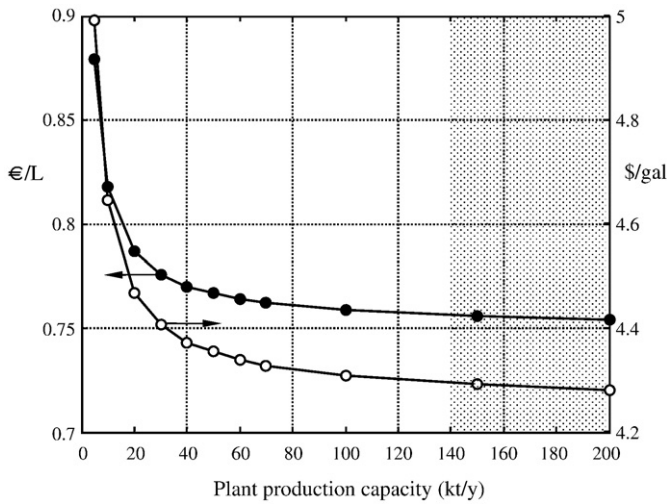
Fig. 3. Unit production cost as a function of plant production capacity.

current price of biodiesel in European markets which is reported to be €0.800/L (≈$4.5/gal).

Until very recently, purified glycerol was considered as a high-value chemical with prices as high as $2/kg used in the manufacture of foods, pharmaceuticals and personal care products [25]. However, the glycerol produced by the biodiesel production plants cannot be absorbed by its traditional market and a glut of glycerol has been created causing sharp decrease in the price of glycerol, which is now estimated to be around $0.1/kg. In all previous studies significant revenues have been attributed to the raw glycerol. As this cannot be the case at the current market situation and until results of research in the raw glycerol utilization are successfully commercialized (such as in the 1-3 propanediol production) no revenues can be attributed to glycerol.

5.4. Cost calculations as a function of production capacity

In Fig. 3 the unit production cost as a function of biodiesel plant production capacity is shown. For production capacity as low as 5 kt/year the unit cost is estimated to be €0.880/L ($5/gal) and decreases sharply as production capacity increases up to 30 kt/year (€0.776/L, $4.4/gal). Further increase in the production capacity results in a less dramatic decrease in the unit cost which approaches €0.760/L ($4.3/gal) as production capacity approaches 140 kt/year. For capacities

above 140 kt/year (dotted area) some of the cost equations used in this study are not valid and the results are presented only to show the tendency. In Fig. 4 the contribution (%) of the raw materials cost in the total production costs is shown. As can be observed from the figure for low production capacities the raw materials cost accounts for 75% of the total production cost and approaches asymptotically 90% as production capacities increase. Finally, in Fig. 5 the fixed capital investment (FCI) required as a function of the production capacity is shown. It is interesting to observe that, in contrary to our experience, the fixed capital investment varies almost linearly as a function of the production capacity. This can be attributed to the almost linear variation of the cost of storage tanks (which dominates the fixed capital investment) with the storage capacity especially for larger capacities.

Based on the observations that we have made an estimation (lower bound) of the unit production cost can be obtained by taking into account only the cost of oil, methanol, utilities and operating labor and all other costs directly related to them:

$$c = \frac{1.05\rho_{\text{Biodiesel}}}{R_{\text{f}}} \left[\frac{MW_{\text{OIL}}\frac{c_{\text{OIL}}}{X_{\text{f}}} + 3MW_{\text{MeOH}}c_{\text{MeOH}}}{3MW_{\text{Biodiesel}}} + 0.0014c_{\text{steam}} \right. \\ \left. + 0.030c_{\text{elec}} + 1.9 \cdot 0.3 \cdot 10^{-6}c_{\text{operator}} \right] \quad (16)$$

where c is the unit production cost of biodiesel in $/L, R_{f} is the fractional recovery of biodiesel produced, $\rho_{\text{Biodiesel}}$ is the density of biodiesel in kg/L, MW denotes the molecular weight in kg/kmol, c_{OIL} the cost of oil in $/kg, c_{MeOH} is the cost of methanol in $/kg, X_{f} is the fractional conversion of oil, c_{steam} is the cost of steam in $/GJ, c_{elec} is the cost of electricity in $/kWh and c_{operator} is the cost of a plant operator in $/year. The unit cost calculated based on Eq. (16) is always lower bound to the actual unit cost and can be achieved at extremely large plant capacities where the fixed capital cost is negligible. By replacing some of the cost coefficients into Eq. (16) the following equation can be obtained:

$$c = \frac{1}{R_{\text{f}}} \left[0.9135\frac{c_{\text{OIL}}}{X_{\text{f}}} + 0.0992c_{\text{MeOH}} + 0.0379 \right] \quad (17)$$

For the fractional recoveries and conversions that have been used in this study Eq. (17) predicts a unit production cost of €0.73/L ($4.14/gal). This value is in agreement with Fig. 3, as expected. It is important to note that this is the minimum unit production cost for plants that are using alkali-catalyzed transesterification and can only be achieved
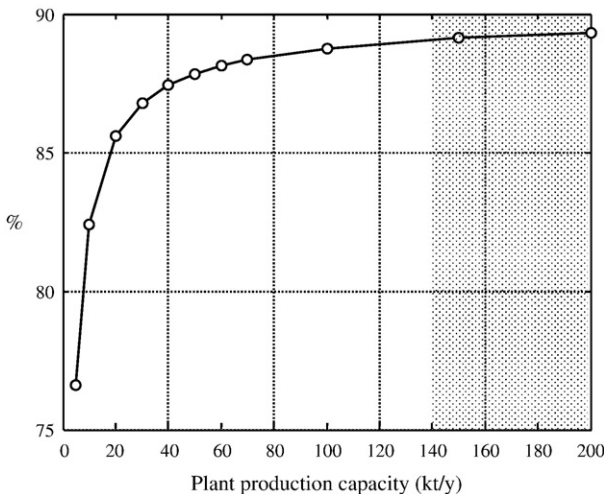

Fig. 4. % of raw materials cost in total production costs as a function of plant production capacity.
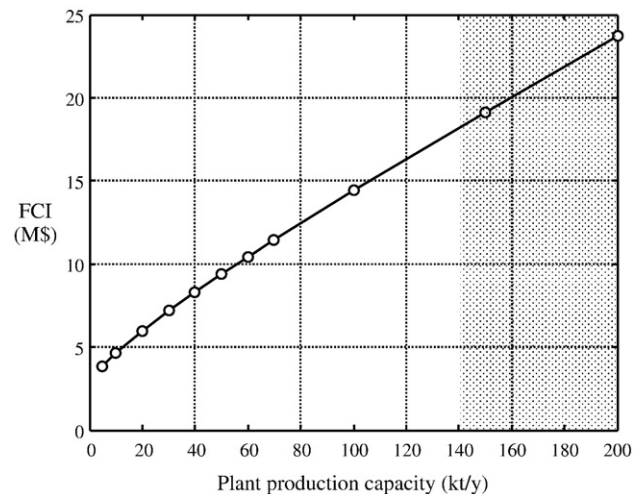

Fig. 5. Fixed capital investment as a function of plant production capacity.

at very large production capacities. They are extremely useful as they can be used in order to quantify unambiguously the potential for exploring the economies of scale when decisions related to the selection of production capacity of new plants are considered.

In order to demonstrate further the usefulness of Eqs. (16) and (17) we will consider the economic study of Hass et al. [8]. They estimate that for a biodiesel production plant producing 33.1 kt/year biodiesel using degummed soy oil with unit price of $0.52/kg the unit production cost (for zero glycerol credit) is $2.123/gal. Application of Eq. (17) gives an estimate of the minimum unit production cost of $2.088/gal. Marchetti et al. [10] estimate a unit production cost of $0.5084/kg for a biodiesel plant producing 36 kt/year using high FFA oils with cost of $0.4/kg. Application of Eq. (17) gives a minimum unit cost of $0.441/kg. If this is increased by 10% to account for the preesterification treatment because of the presence of FFA then the minimum unit production cost estimation will be $0.486/kg.

6. Profitability indicators

The calculation of the profitability of a chemical process is performed using methods such as the payback period (PBP), the net present worth (NPW), the return on investment (ROI) and the discounted cash flow rate of return (DCFR). In this work the ROI and the DCFR indices have been calculated as a function of plant capacity and are shown in Fig. 6. In Fig. 7 the NPW for the cases of interest rate of 10% and 20% is shown. The total capital investment as a function of plant capacity is also shown in the same figure. The main assumptions that have been used in the underlying calculations are the following:

- plant construction will last for 2 years, 40% of the fixed capital will be required in the first year
- working capital can be estimated as the capital required to cover all variable costs for 1 month of operation at full capacity
- plant operation for 10 years with 50% capacity achieved in the first year, 90% in the second year and 100% capacity achieved in the remaining years
- salvage value of all equipment and land can be neglected
- depreciation is calculated using the modified accelerated cost recovery system (MARCS) with a recovery period of 7 years
- the selling price of biodiesel is €0.800/L
- taxation rate at 25% of the gross profits (depreciation included)
- zero glycerol revenues

As can be seen from Figs. 6 and 7 capacities below approximately 15 kt/year should be avoided as the plant operation cannot be
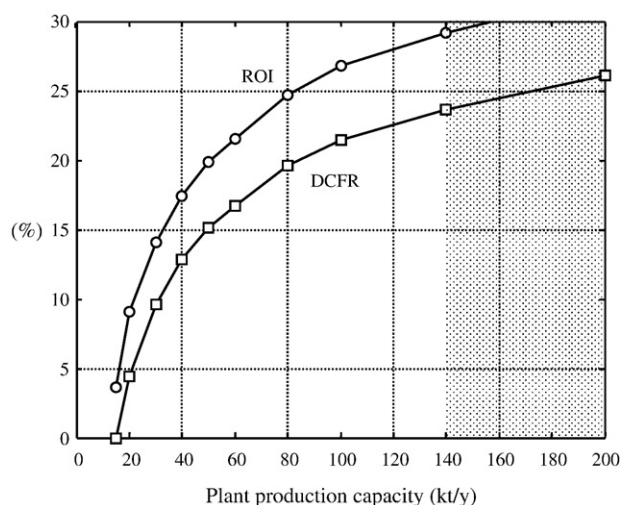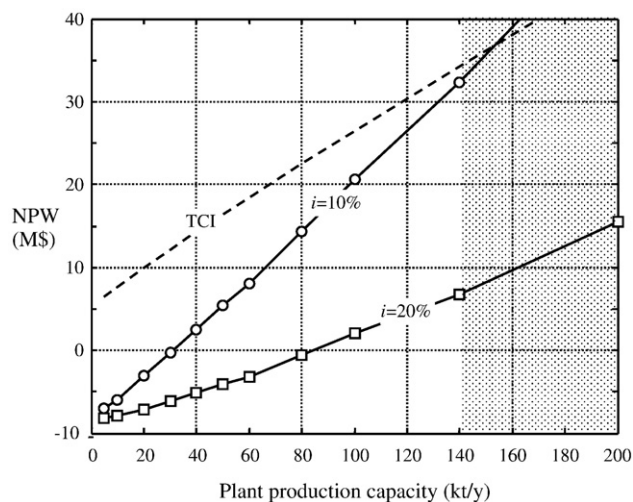
Fig. 7. NPW and TCI as a function of plant production capacity.

profitable. Plants with installed capacities between 15 kt/year and 30 kt/year can be considered as having a high level of risk as the DCFR is below 10%. Plants with installed capacities between 30 kt/year and 50 kt/year can be considered as having medium risk as the DCFR lies in the area of 10–15%.

According to the results shown in Figs. 6 and 7 the minimum acceptable capacity is that of 50 kt/year while capacities larger than 80 kt/year should be considered in order to reduce the risk to an acceptable level.

7. Conclusions

In this work the economics of biodiesel production plants that use classical alkali-catalyzed transesterification is studied for the first time as a function of plant installed capacity. Estimations of the unit production cost and fixed capital investment for a wide range of plant production capacities have been calculated in detail. Based on the observation that the raw materials cost accounts for 75% of the total production costs for small plants and can increase to 90% for large plants a simplified equation has been proposed for the calculation of a lower bound on the unit biodiesel production cost. The main conclusions are that plants with capacities less that 15 kt/year should be avoided as the plant operation cannot be profitable while only plants with capacities greater than 50 kt/year–80 kt/year can be feasible from the economic point of view. These conclusions have some serious implications as the current mean installed capacities are of the order of 40 kt/year. Furthermore, the results of this study justify the current tendency to build biodiesel plants with capacities significantly larger than the currently mean installed capacity. Small plants that have already been installed with capacities in the range of 10–50 kt/year will be extremely difficult to remain competitive. Needless to say that the competition for the control of the prices of raw materials and especially oil will be fierce as the larger plants will need to extend their quest for raw materials to extended areas. This is particularly true for countries like Greece where, under the current market situation and biodiesel consumption [26], the existence of only a few of large biodiesel production plants can be feasible in the long term.

Acknowledgement

Fig. 6. DCFR and ROI as a function of plant production capacity.

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