International Journal of ChemTech Research
CODEN (USA): IJCRGG ISSN: 0974-4290
Vol.8, No.9, pp 297-304, 2015

Biodiesel production from Chlorella Sp: Process Design and Preliminary Economic Evaluation

N. Mohammady[1], S.I. Hawash[2], K.M. El-Khatib[2], M.I. El-Galad[3], G. El Diwani[2]

[1]Botany and Microbiology Dept, Faculty of Science, Alexandria University
[2]Chemical Engineering and Pilot Plant Dept, National Research Centre
[3]Chemical Engineering Dept, Faculty of Engineering, Cairo University

Abstract: Biodiesel has become an attractive diesel fuel substitute to its environmental benefits since it can be made from renewable resource. A technological assessment of biodiesel production from chlorella marine was carried out to evaluate their technical benefits and limitations. The main issue with algal oil is the high level of free fatty acids found in it. Economic analysis is a powerful tool that can be used to both evaluate the production cost of algae bio-fuels and help to identify the major factors which contribute most to the production cost – thereby helping to focus future engineering research. This investigation examines how economic assessment has been used to estimate the future cost of transport fuels produced from algal biomass. The acid-catalyzed process using algal oil proved to be technically feasible with less complexity than the alkali-catalyzed process, thereby making it a competitive alternative to commercial biodiesel production by the alkali-catalyzed process. It was found that total capital investment for 10,000 tons biodiesel per year is18, 270, 783 $. Total manufacturing cost was $29, 386, 124 and the return of investment 46,792%.
Keywords: Biodiesel production, microalgae, Chlorella marina, Process design, Economic analysis.

1. Introduction:

Biodiesel is a renewable and sustainable substitute of the diesel fuel traditionally obtained from petroleum. The main benefit of this fuel is related to the environmental advantages compared to traditional fuel [1].It is derived from a renewable, domestic resource, thereby relieving reliance on petroleum fuel imports. It is biodegradable and non-toxic. Compared to petroleum-based diesel, biodiesel has a more favorable combustion emission profile[2,3]. Transesterification reactions can be alkali-catalyzed, acid-catalyzed or enzyme-catalyzed. The first two types have received the greatest attention[4]. Microalgae are characterized by higher biomass production and faster growth compared to other energy crops. Moreover, algal oil is an interesting alternative to popular feedstock on nutritive crops[5], Since it does not compete with these traditional foods [6].

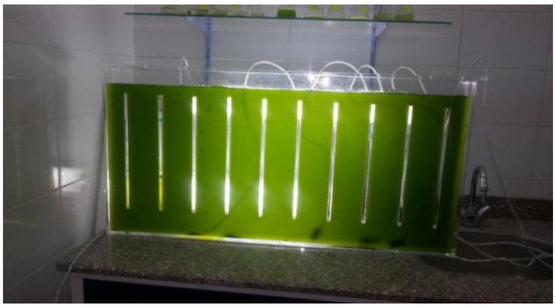

Figure 1: Newly installed Flat Panel Photo-bioreactor

2. Experimental Work:

At laboratory scale chlorella marina was cultivated in flat panel bioreactor was shown in figure (1). Figure (2) illustrates Chlorella sp. under microscopy

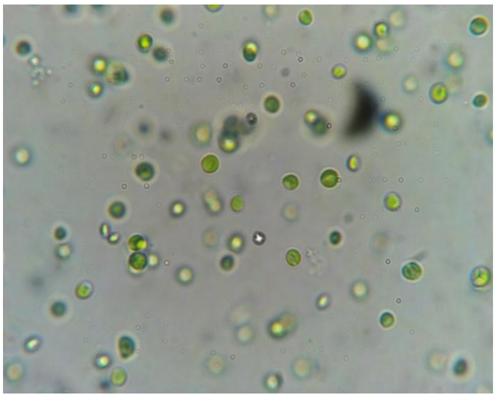

Figure 2: *Chlorella* sp. Under Light Microscopy

Then Chlorella inoculums was transferred to open pond chamber as shown in figure(3).

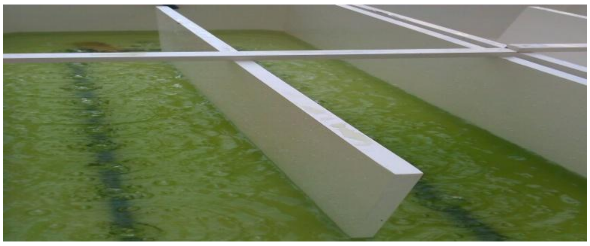

Figure 3: An open pond with *Chlorella* sp

Different conditions for chlorella marina growth were studied and according to laboratory results the growth of Chlorella sp cultivated under different conditions has increased approximately 3-times (from 1.3 to 4 g/l) at:

- pH: 8.5
- Salinity: 45 g/l
- Nitrate: 4 g/l
- Dilution: 30%

By characterizing produced Chlorella sp. it was found that

- Lipid content: 800 mg/l
- Acid Value: 0.449
- Saponification Value: 50.5
- M.W: 336.26; calculated as an average of fatty acid profile molecular weights.

By-Products of delipidated cell compared to the complete one **(expressed in %)**

- Protein content of complete cell:118.50 mg/l
- Protein content of delipidated cell:100.50 mg/l
- Protein in the remnants: 84.8% of the complete cell.
- Carbohydrate content of complete cell:20.6mg/l
- Carbohydrate content of delipidated cell:15.5mg/l
- Carbohydrate in the remnants:75.0 % of the complete cell.

But the above by-products are not taken in consideration for economic evaluation in this investigation.

3. Results and discussion:

3.1. Process simulation:

To assess the commercial feasibilities of the proposed process, complete process simulation was first carried out. Despite some expected differences between process simulation results and actual process operation, most current simulation software's can provide reliable information on process operation because of their comprehensive thermodynamic packages, vast component libraries and advanced calculation techniques. The process simulation software, Aspen HYSYS V7.0 developed by AspenTech Inc., was used in this study.

The procedures for process simulation mainly involve defining 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, sulfuric acid, sodium hydroxide and water, is available in the HYSYS^{TM} component library. Regarding the oil feedstock, triolein ($C_{57}H_{104}O_6$) was considered as the oil obtained from microalgae in this process because it is the most common triglyceride in microalgae's oil. Accordingly, methyl oleate ($C_{19}H_{36}O_2$) was taken as the resulting biodiesel product and its properties were available in the HYSYS^{TM} component library. For those components not available in the library, such as sodium sulfate, they were defined using ''the Hypo Manager'' tool in HYSYS^{TM}. Other physical properties, such as critical temperature, pressure and volume, were estimated by HYSYS^{TM}.

Due to the presence of methanol which is a highly polar component, both the non-random two liquid (NRTL) and universal quasi-chemical (UNIQUAC) thermodynamic/activity models were recommended to predict the activity coefficients of the components in a liquid phase (AspenTech Inc. 2005). Detailed descriptions of these models were provided by Gess et al. (1991). The NRTL model was used in this study.

The industrial scale production of biodiesel in Egypt is in its first steps. The plants are built with small or medium capacities. This study is passed on 10,000 ton/year biodiesel.

3.2. Process Design

Stainless steel (type 316) should be used when sulfuric acid concentrations were below 5 wt.% or above 85 wt.% and temperatures were below the boiling point of the sulfuric acid solution[7]. Stainless steel 316 was indicated to have an acceptable corrosion rate (less than 0.5 mm/yr) for 5 wt.% sulfuric acid under 100 ^oC[8]. Depend on these basics and the use of sulfuric acid in our process, stainless steel 316 will be used for almost all equipment as material of construction.

3.2.1. Process description

3.2.1.1. Cultivation

As shown in figure (4) the pre-prepared microalgae are being fed into open ponds to be cultivated. It took days to reach the optimum growth of microalgae. The nutrients are fed into the ponds to facilitate the growth of the microalgae; and a stream of water to make up the loss from evaporation. After that, the grown microalgae media is directed to harvesting facility.

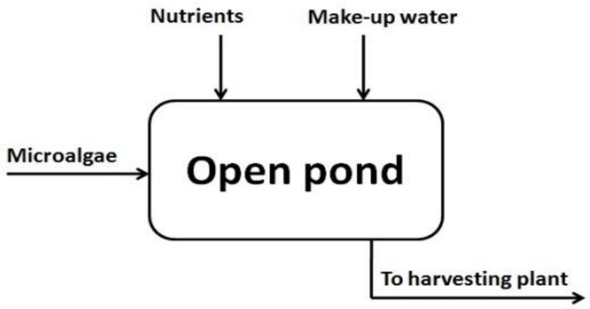

Figure 4: Cultivation flow sheet.

3.2.1.2. Harvesting

The water stream from the open pond contains 99.83% water; this water amount must be removed to ease the extraction process figure (5). The remove of water is done by two steps: first by autoflocculators; then second by centrifuge. After that, a drying step by sand bed dryers is used to dry the microalgae nearly to 78% dry weight basis (DWB).

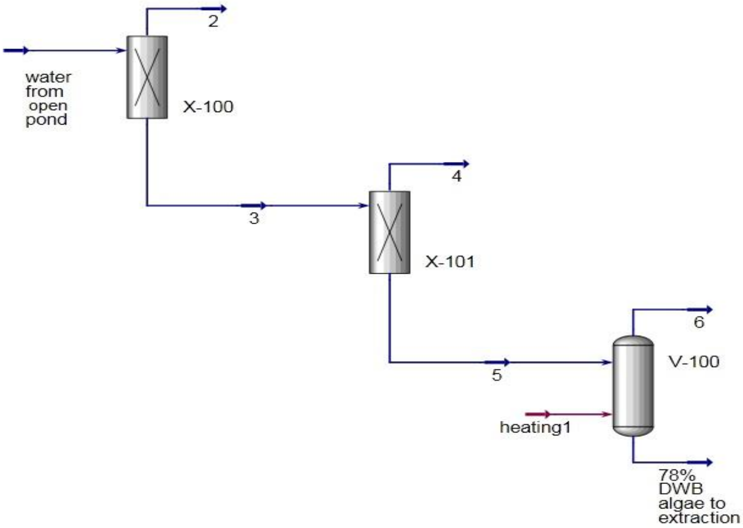

Figure 5: Harvesting flow sheet; where (X-100) autoflocculation, (X-101) centrifuges and (V-100) sand bed dryers.

3.2.1.3. Extraction

Solvent extraction is the most feasible technique to extract oil from microalgae. Pure hexane was used in the process. After extraction, the solvent/oil mixture is separated from the algae cake in separator X-102. The mixture is directed to an evaporator V-101 to recover hexane with efficiency 99.9%. The extracted oil with 99.985 wt% is then used to get biodiesel using transesterification reaction as shown in figure (6).

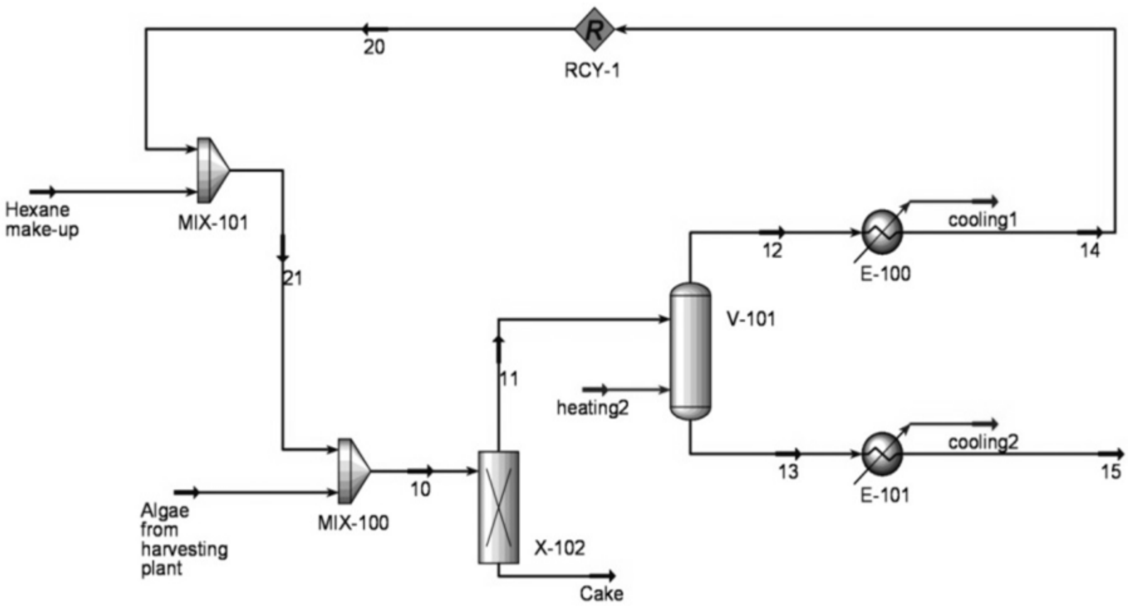

Figure 6: Extraction flow sheet; where (MIX-100) microalgae/solvent mixer, (MIX-101) recycled/make-up solvent mixer, (X-102) settling tank, (V-101) hexane evaporator, (E-100) recycled hexane cooler and (E-101) oil cooler.

3.2.1.4. Transesterification

A continuous transesterification process flow sheet using the extracted oil was developed as shown in figure (7). The transesterification reaction was carried out at 80 ^{o}C, 400 kPa. The fresh methanol stream, the recycled methanol stream 1201 and the H_2SO_4 stream were mixed before being pumped into transesterification reactor CRV-100 by pump P-100. In CRV-100, 97% conversion of triglycerides to methyl esters was achieved. After the transesterification reaction, stream 106 was forwarded to distillation tower T-100 to remove excess methanol.

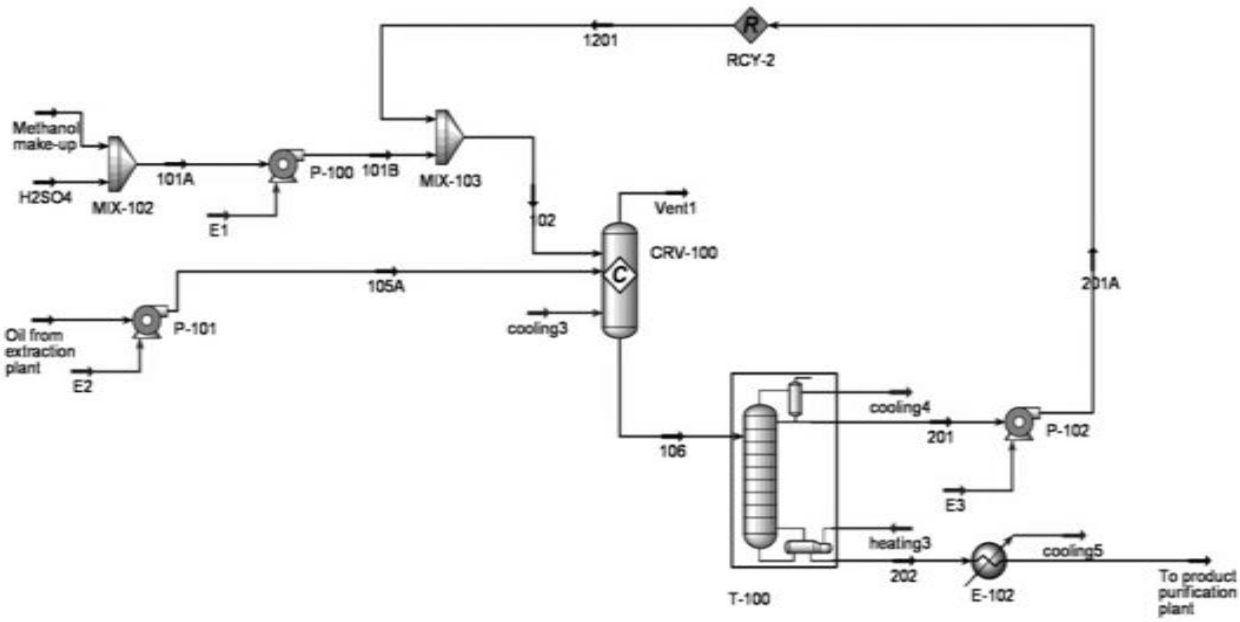

Figure 7: Transesterification flow sheet; where (MIX-102) make-up alcohol/catslyst mixer, (MIX-103) recycled alcohol/catalyst and fresh alcohol mixer, (P-100) alcohol and catalyst pump, (P-101) oil pump, (P-102) recycled alcohol pump, (CRV-100) transesterification reactor, (T-100) methanol recovery distillation tower and (E-102) cooler.

3.2.1.5. Product purification

After methanol recovery, the bottom stream was directed to reactor CRV-200 to neutralize the H_2SO_4 catalyst. In R-200, sulfuric acid was removed by adding calcium oxide (100% purity). The resulting $CaSO_4$ was removed in gravity separator X-103.

To separate the FAME from the glycerol, methanol and catalyst, a water washing column (T-102) was used. The FAME in stream 203C was separated from the glycerol, methanol and catalyst by adding water (25 ^{0}C). All of the glycerol remained in the bottom stream 302.

In order to obtain a final products, both streams 301 and 302 was directed to distillation towers T-103 and T-104 respectively; to obtain biodiesel (99.6% pure) and glycerol (86% pure) respectively see figure 8.

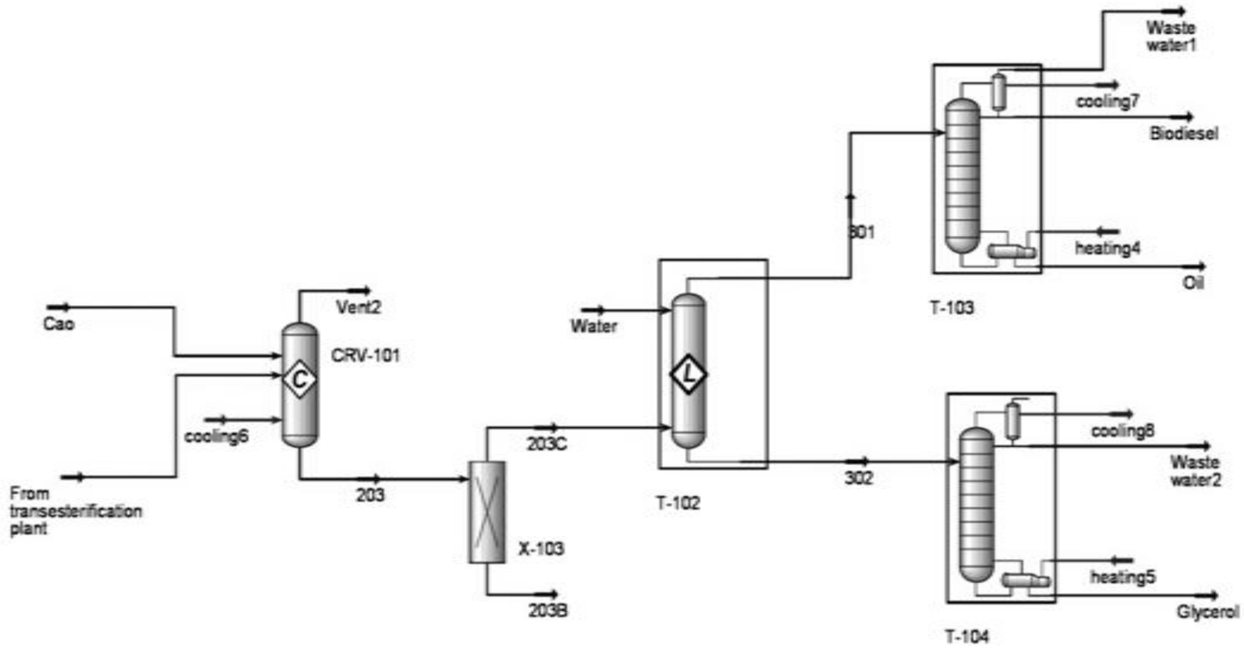

Figure 8: Product purification flow sheet; where (CRV-101) neutralization tank, (X-103) settling tank, (T-102) water washing tower, (T-103) biodiesel purification distillation tower and (T-104) glycerol purification distillation tower.

3.2.2. Equipment sizing

Process equipment was sized according to principles outlined in the literature[9-11].

3.2.2.1 Reactor vessels

Reactors were sized for continuous operation by dividing the residence time requirement by the feed flow rate for each process. Residence times were obtained from experiments and they were: 4 hrs and 10 min for the transesterification reactor and neutralization reactor, respectively. The vessels were specified to have an aspect ratio of 1:1.5.

3.2.2.2. Distillation column

Distillation column diameter was sized using the Souders-Brown equation. While tray tower height was calculated by multiplying the number of actual stages by the tray spacing, and then increasing the result by 20% to provide height for the condenser and reboiler.

3.2.2.3. Gravity separators and mixers

The gravity separators and mixers were designed as vertical process vessels with an aspect ratio of 1:1.5. They were sized to allow for continuous operation, with a residence time of 1h.

3.2.2.4. Evaporators and heat exchangers

The evaporator and heat exchangers were designed by simple energy balance equation; from which the heat transfer area was calculated. The evaporators were designed as falling film type; while the heat exchangers were designed as shell and tube type with fixed head.

3.3. Operational costs

The technical evaluation is not the only factor to evaluate a project; as there are other factors like economical, environmental and social factors. The economic performance is an important factor in assessing process viability. It checks the project's profitability, either it will lose money; or it will earn money. The economic performance of a biodiesel plant (e.g., fixed capital cost, total manufacturing cost, and the break-even price of biodiesel) can be determined once certain factors are identified, such as plant capacity, process technology, raw material cost and chemical costs.

This economic evaluation was based on the following assumptions: (1) Operating hours for the biodiesel plant was assumed to be 8000 h/year. (2) In the simulation, pump efficiency was assumed to be 70%. This was used to determine the pump shaft power. Every pump is considered to have a spare one. (3) Low pressure steam was used as the heating media. Water was the cooling medium. Their specifications and prices are listed in Table 1. (4) All costs shown are in US$. Equipment prices were updated from available 2001 or 2007 to 2013 values using the Chemical Engineering Plant Index, where I_{2013} = 567.3, I_{2001}= 397 and I_{2007} = 525.4[12,13]. (5) All chemical costs including raw materials, catalysts and products are given in Table 1 according to Egyptian local market prices.

Table 1: Costs of raw materials, products and utilities used in the process.

Raw materials	
Hexane ($/ton)	791
Methanol ($/ton)	440
H_2SO_4 ($/ton)	2398
CaO ($/ton)	150
Water ($/ton)	0.007
Products	
Biodiesel ($/ton)	1235
Cake ($/ton)	300
Glycerol ($/ton)	1200
Utilities	
LPS ($/ton)	6.8
HPS ($/ton)	10
Cooling water ($/m^3)	0.007
Electricity ($/kw.hr)	0.04

According to the definition of capital cost estimation[12] the economic estimation in this article is classified as a ‘‘study estimate’’. It is based on the development of a process flow diagram and rough sizing of major process equipment. No further information, such as a layout plot, process instrumentation diagram or piping and instrumentation requirements, were considered. This study estimate had a range of expected accuracy from +30% to -20%[12]. Thus, results from such a preliminary evaluation may not accurately reflect the final profitability of a chemical plant but can be used as a tool for comparison of several process alternatives[12]. The economic assessment was developed by the literature outlined by[14].

3.3.1. Total capital investment (TCI)

The total capital investment is divided into fixed and working capital investments. The fixed capital investment (FCI) is defined as the investment needed to make the plant ready for start up and it includes the costs of equipment, installation, piping, instrumentation, electrical, building, utilities, storage, site development, ancillary buildings, design, contractor's fee and contingency. While for the working capital investment (WCI), it is defined as the investment needed to run the plant for 3 months, where in these 3 months, all the variables are adjusted till the plant is ready for real production. The total capital investment beside to the purchased costs of main equipment is found **$18,270,783**.

3.3.2. Total manufacturing cost (TMC)

In order to sell a product and decide its price, its manufacturing cost must be known so as to put a profit on it and decide its selling price. The manufacturing cost divides into direct and indirect manufacturing costs. The direct manufacturing cost (DMC) includes costs of raw materials, miscellaneous, utilities, shipping and packaging, labor, supervision, plant overhead, depreciation, interest, insurance, rent, royalties and maintenance. For the labor cost, it was estimated as $493,458; and it was calculated by assuming that an operator worked 48 weeks/year and there were three 8-h shifts per day for the continuous plant.

The indirect manufacturing cost (IDMC) includes the sales, distribution, general overheads, research and development; and It equals to 25% of the DMC.

Total manufacturing cost is calculated and found **$29,386,124.**

4. Conclusions

To get Internal rate of Return of Investment (ROI) the net profit must be calculated. The net profit is the money gained or loosed resulting from revenues after subtracting all associated costs. It can be calculated by subtracting the TMC from the income from sales. The main product of this project is the biodiesel which costs 1235$/ton. The glycerol and algae cake are by-products for this process but they are valuable too, as they cost 1200$/ton and 300$/ton respectively and produced in a high rate. The total products' sales cost equals to $38,729,840. Depreciation is an income tax deduction that allows a taxpayer to recover the cost or other basis of certain property. It is assumed to be 5% of the FCI.

By using equation (1), the ROI will equal to 46.792%. The ROI is used to check the feasibility of a project by comparing it to the minimum acceptable rate of return (MARR). Assume MARR is 20%. Observations for this project results in that ROI > MARR; so, the project is feasible.

$$ROI\% = Profit / (FCI + Dep.) \qquad (1)$$

The break-even point is the point at which the product stops costing money to produce and sell, and starts to generate a profit for the company. At break-even point, the production cost equals to income from sales. So, for this process, the break-even point is at 714 days of production.

References

1. Srivastava A, R Prasad. Triglycerides- based diesel fuels, Renew. Sustain. Energy Rev2000; (4): 111-133.
2. Korbitz W. Biodiesel production in Europe and North American. an encouraging prospect, Renew. Energy 1999; (16): 1078-1083.
3. Agarwal A K, Das L.M. Biodiesel development and characterization for use as a fuel in compression ignition engines, J. Eng. Gas Turbines Power 2001; (123): 440-447.
4. Zhang Y, M ADube , D D, McLen M Kates .Biodiesel production from waste cooking oil: 1. Process design and technological assessment**,** Bioresource Technology 2003;(89): 1-16.
5. Minowa T, Yokoyama S Y, Kishimoto M and Okakurat T. Oil production from algal cells of Dunaliella tertiolecta by direct thermochemical liquefaction, Fuel 1995 ; (74): 1735-1738.
6. Chen P, Min M, Chen Y, Wang L, Li Y, Chen Q, Wang C, Wang Y, Wang X, Cheng Y, Deng S, Hennessy K, Lin X, Liu Y, Wang Y, Martinez B and Ruan R. Review of the biological and engineering aspects of algae to fuels approach, Int. J.Agric. Biol. Eng. 2009 ; (2): 1-30.
7. Green D and R Perry. *Perry's Chemical Engineers' Handbook* 8th ed. 2007; New York, USA: McGraw-Hill.
8. Davis JR. *Corrosion: Understanding the Basics* 2000; Ohio, USA: ASM International.
9. Kern DQ.*Process Heat Transfer*, 1950;New York, USA: McGraw-Hill.
10. McCabe WL, Smith JC and Harriott P.*Unit Operations of Chemical Engineering,* 5th ed. 1993; New York, USA: McGraw-Hill.
11. SinnottRK,*Coulson and Richardson's. Chemical Engineering, 1999;(6): Chemical Engineering Design* 3rd ed., London, UK: Butterworth-Heinemann.
12. Turton R et al. *Analysis*, *Synthesis and Design of Chemical Processes*, 3rd ed.2009; New Jersey, USA: Prentice Hall.
13. Economic Indicators, Chemical Engineering 2013 (October); Available at: www.che.com.
14. Peters Mand Timmerhaus K. *Plant Design and Economics for Chemical Engineers*, 1991; 4th ed., Singapore: McGraw-Hill, Inc.Accessed October 31, 2014; Available at: http://www.amazon.com/Plant-Design-Economics-Chemical-Engineers/dp/0072392665.

*****