

# Flexible Modular Process Design for Enzymatic Biodiesel Production

Catarina Sanches Seita

Instituto Superior Técnico, Lisbon, Portugal

June 2013

---

**Abstract:** Nowadays feedstock cost constitutes one of the biggest contributors for overall biodiesel production costs, which causes small margins for profit of biodiesel industrial plants. Hence, it is important to develop processes with higher flexibility, regarding quality and availability of feedstocks for biodiesel production. This project studied the concept of a generic process plant where a range of feedstocks can be used in the same plant, via flexible modular process design. The combination of Crude Palm Oil (CPO) and Palm Fatty Acid Distillate (PFAD) in a flexible enzyme-catalyzed process has been assessed. Using a process simulation software – SuperPro Designer® - the main technical and economical differences in the process performance for each feedstock have been evaluated. Simulation results show that the differences in PFAD's critical parameters, namely its composition and its melting point, cause a significant reduction in the volumetric efficiency of the reactor when using this feedstock. Annual biodiesel production is approximately 42% lower if using PFAD in the process and annual glycerol production is approximately 90% lower.

**Keywords:** Flexible process; Process simulation; Enzymatic biodiesel; crude palm oil; palm fatty acid distillate (PFAD)

---

## 1. Introduction

Over the past decades there was an increasing growth in the biodiesel industry. One of the main reasons for that is the need to develop alternatives to fossil fuels. The need to decrease the dependency on foreign energy supply from fossil fuel resources led to governmental and financial incentives which boosted biodiesel production <sup>[1], [2]</sup>.

Biodiesel has a lot of environmental and technical benefits. Regarding the environmental benefits, it has reduced emissions of particles, sulfur, carbon monoxide and hydrocarbons <sup>[2]</sup>; reduced emission of greenhouse gas <sup>[1]</sup>; it is biodegradable <sup>[3]</sup> and its use in place of conventional fuels would reduce the global warming effect <sup>[2]</sup>. Technically speaking, it has a higher flash

point than petroleum diesel, which makes it easier to handle, transport and store <sup>[4]</sup>; it shows an excellent lubricity which reduces engine wear and therefore prolongs engine life <sup>[4]</sup> and finally, it can be used in the existing transport sector, in contrast with other technologies such as hydrogen or electrical vehicles which would require major changes in the sector <sup>[1]</sup>.

However, it also has some drawbacks, namely a lower energy content than diesel fuel <sup>[3]</sup> and the fact that biodiesel is more likely to suffer oxidation than petroleum diesel.

Currently the high cost of vegetable oils, which are the main type of feedstock used for biodiesel production, is jeopardizing the

economic viability of many production plants. Several reports have been published with an estimation of biodiesel production costs and in all of them feedstock cost constitutes up to 70-95 % of total biodiesel production costs <sup>[5], [6]</sup>. Therefore, it is important to develop technologies that allow the use of low-quality feedstocks, such as waste oils, in a way that is economically viable and profitable. Besides providing a reduction in biodiesel production cost, using waste oils would avoid the food vs. fuel debate.

The design of a multi-feedstock plant that could accept several types of feedstocks, in order to accommodate variations in feedstock quality and availability, would improve the overall economy of biodiesel industry.

There are a number of questions that arise when facing the challenge of designing a multi-feedstock plant. The necessary unit operations depend on the type of feedstock; the reaction conditions as well as the process equipment might also differ according to the feedstock, so it is an enormous task to combine all these variables into a single generic process.

This project was focused on a case study- the combination of Crude Palm Oil (CPO) and Palm Fatty Acid Distillate (PFAD) in a flexible enzyme-catalyzed process. PFAD is a by-product of physical refining of CPO and therefore considerably cheaper than CPO.

An enzymatic catalyst is an interesting option for a flexible process as it shows several advantages - mild operation conditions, glycerol with a higher quality-grade, and more importantly, it has the ability to esterify both FFA and TAG into FAME, which provides a higher flexibility regarding the feedstocks quality <sup>[1]</sup>.

## 2. Materials and methods

To support the process design, an extensive literature review (about feedstocks characteristics; catalysts options; unit operations and operation conditions; etc.) was made. The critical parameters (melting point; FFA, water and phospholipids content) of CPO and PFAD were analyzed to determine which unit operations are necessary for the feedstocks pre-treatment.

A process flowsheet was designed for CPO. The process was designed in a batch mode and with an annual capacity of 100,000 tons. In order to have a high flexibility the process was designed in a modular way, in which different feedstocks might have a different combination of unit operations throughout the entire process.

The designed process was based on the BioFAME process from Novozymes A/S (patent pending), shown in Figure 1 <sup>[7]</sup>.

The process includes four sections- the feedstock pre-treatment section, which includes the feedstock melting, filtration to remove impurities and degumming; the reaction section; the biodiesel purification section which includes all the downstream operations (glycerol separation, methanol removal, alkaline washing, water washing and drying) and the glycerol refining section which includes a distillation column to remove the remaining water and alcohol.

Based on information from the supplier (Novozymes A/S) it was assumed that the enzyme loses activity after 5 batches. So in order to reuse the enzyme during 5 batches, 80 %w/w of the glycerol stream (after separation from biodiesel) is recirculated back to the reactor. The remaining 20% of the water/glycerol stream is sent to the glycerol refining section. The enzymatic catalyst is not separated from the glycerol stream, but with the high temperatures applied in the distillation the enzyme will denature.

The process was then adapted to PFAD but no changes were made in the process equipment or in the operation conditions, with exception of the transesterification time in the reactor. The enzyme used is Callera™ Trans L (commercialized by Novozymes A/S) which is a lipase from *Thermomyces lanuginosa* (TLL). TLL has a high specificity towards TAG. CPO has a high content of TAG whereas PFAD has a high content of FFA (80-95%). Therefore it was assumed that the reaction would be faster for CPO (16 hours vs. 20 hours for PFAD).

The high melting point of PFAD (49°C) imposes a big challenge when adapting the process. The reaction temperature is 39 °C

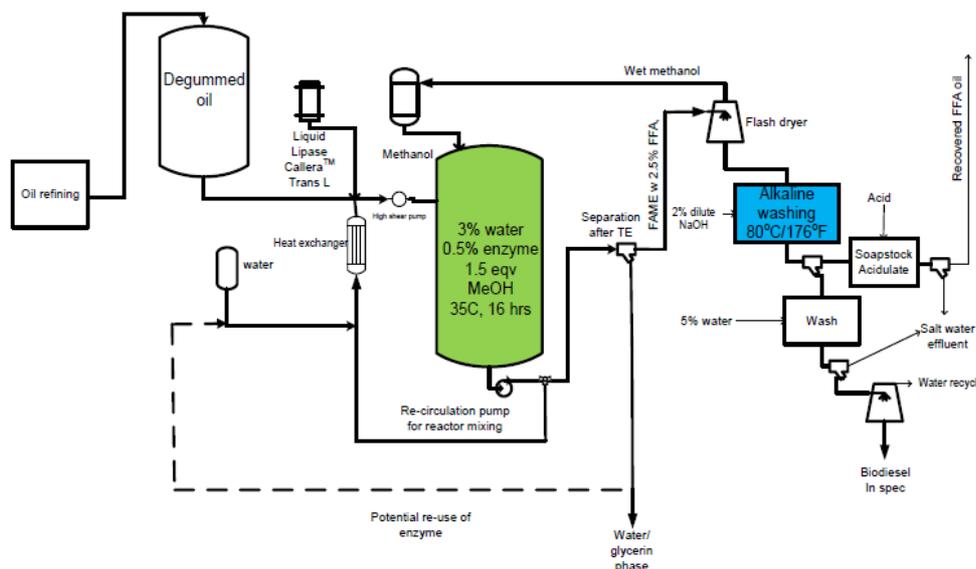


Figure 1 BioFAME process layout [7].

and it cannot be higher than this value as enzymes might suffer denaturation.

The strategy adopted to decrease PFAD melting point is to recirculate 50 %w/w of the biodiesel stream, (after separation from glycerol) blend it with PFAD and it was assumed that this procedure would decrease the melting point to the desired temperature, 39 °C (assumption based on experimental work currently being done in PROCESS group from DTU).

To assess the differences in the process performance for both feedstocks, two process simulations were carried out. The simulation software chosen was SuperPro Designer®, version 8.5, developed and commercialized by Intelligen, Inc. The procedures for process simulation involve defining the chemical components, choosing the right operation units and operation conditions (temperature, pressure, flowrate, etc.) and design the process flowsheet. The operation conditions for the several units in the process are shown in Table 2.

### 3. Results and discussion

#### 3.1 Process design

The critical parameters of CPO and PFAD are shown in Table 1.

Table 1 Critical parameters for CPO and PFAD.

Critical parameters	CPO	PFAD
FFA (wt%)	5 <sup>[8]</sup>	85 <sup>[8]</sup>
Melting point (°C)	35 <sup>[9]</sup>	49 <sup>[8]</sup>
Phosphorus content (ppm)	5-130 <sup>[10]</sup>	1.14 <sup>[8]</sup>
Water (wt%)	~0.02 <sup>[11]</sup>	~0.87 <sup>[8]</sup>

The main differences between the critical parameters of both feedstocks concern FFA and phosphorus content, and the melting point. The difference in water content is not significant and considering that the process is enzymatic, there is no need to dry the feedstock since enzymes are able to handle feedstocks with some water.

Regarding the high FFA content, since the process is enzymatic there is no need of a pre-esterification step as in a conventional alkaline process because enzymes are able to esterify FFA into FAME.

The difference in phosphorus content indicates that CPO needs a degumming step whereas PFAS does not need it. Finally, regarding the melting point, both feedstocks have a melting point higher than room temperature so it is necessary to melt the feedstock in the beginning of the process. As previously discussed, the strategy adopted to decrease PFAD melting point to 39 °C is to recirculate 50 wt% of the FAME and blend it with PFAD.

Table 2 Operation conditions.

Operation	T (°C)	Pressure	Conditions	Reference
<b>Feedstock melting</b>	CPO: 39 PFAD: 50	1.0 atm	-	-
<b>Filtration</b>	39	1.0 atm	Cake dryness: 5%; Maximum cake thickness: 15 cm	-
<b>Degumming</b>	100	1.0 atm	0.075 wt% of a 85 %w/w solution of H <sub>3</sub> PO <sub>4</sub> (with respect to oil); 1.25 wt% of bleaching earth (with respect to oil); Intense mixing (2kW/m <sup>3</sup> ) for 30 minutes; Reaction extent: 100%	[12]
<b>Transesterification</b>	39	1.0 atm	5 %w/w of water; 0.75 %w/w of enzyme Callera™ Trans L; 1.5 eq. MeOH; Intense mixing (2 kW/m <sup>3</sup> ); Reaction time: CPO- 16 hours; PFAD- 20 hours, Transesterification reaction extent: 95%; TAG hydrolysis extent: 95%; Extent of esterification of FFA defined in order to have 2.5 %w/w FFA (with respect to oil) on the outlet stream of the reactor.	Experimental work done in PROCESS group from DTU
<b>Glycerol separation</b>	39	1.0 atm	Separation between oil and glycerol phases: 100%; TAG, FAME and FFA partitioning: 100% in the oil phase; MeOH partitioning: 85% in the glycerol/water phase and 15% in the oil phase; Water partitioning: 99.7% in the glycerol/water phase and 0.3% in the oil phase.	[13]
<b>Methanol removal</b>	120	1.0 atm	MeOH evaporation: 99%; Water evaporation: 90%	Previous projects from PROCESS group from DTU
<b>Alkaline washing</b>	41.8	1.0 atm	10% excess of NaOH (solution of 2 %w/w dilute NaOH)	[7]
<b>Water washing</b>	40.2	1.0 atm	5 %w/w of water (with respect to oil phase)	[7]
<b>Drying</b>	40.42	0.2 bar	-	[5]
<b>Glycerol refining</b>	Condenser: 100°C Reboiler: 127.74 °C	1.0 atm	Reflux ratio:0.123; Theoretical stages: 9.6; Stage efficiency of 80%.	[5]

A simplified process layout is shown in Figure 2.

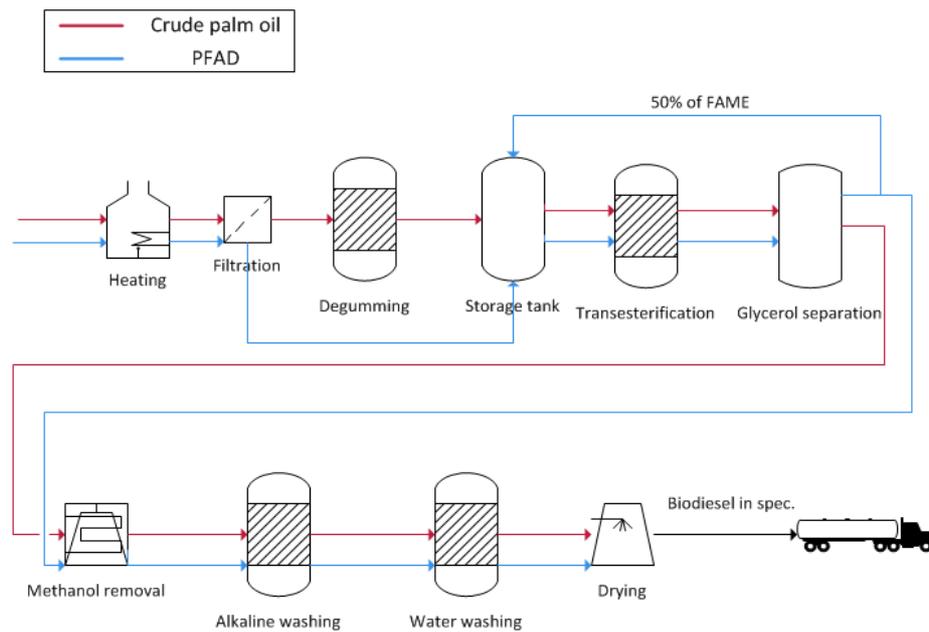


Figure 2 Simplified process flowsheet.

### 3.2 Process performance evaluation for CPO case

It was considered that the batch time is 24 hours (16 hours for the reaction plus 8 hours to fill and empty the reactor), which means the batch capacity must be approximately 303 tons of biodiesel per day. The feedstock pre-treatment and biodiesel purification sections also work during 24 hours - while the reactor is occupied converting CPO into biodiesel and glycerol, the feedstock pre-treatment section is treating CPO for the following batch and the biodiesel purification section is purifying biodiesel from the previous batch.

The recirculation of 80 %w/w of the glycerol/water phase back to the reactor will lead to an accumulation of glycerol in the reactor in the first batches. But this recirculation of glycerol will stabilize and the process will reach a "pseudo steady-state". In this "pseudo steady-state" the consumption of methanol, water and enzyme will be lower - it corresponds to the difference between the amount necessary (defined by the operation conditions) and the amount that is recirculated.

In Table 3 is shown the materials consumption for the entire process in "pseudo steady-state".

Table 3 Materials consumption for the process running in "pseudo steady-state" in CPO case.

Component	Unit operation	kg/batch	ton/year
Crude palm oil	-	$3.16 \times 10^5$	$1.04 \times 10^5$
Methanol	Transesterification	$3.84 \times 10^4$	$1.27 \times 10^4$
Liquid lipase	Transesterification	$4.64 \times 10^2$	$1.54 \times 10^2$
Water	Transesterification	$2.65 \times 10^3$	$8.75 \times 10^2$
	Alkaline washing	$6.00 \times 10^4$	$1.98 \times 10^4$
	Water washing	$1.55 \times 10^4$	$5.12 \times 10^3$
	TOTAL	$7.81 \times 10^4$	$2.58 \times 10^4$
NaOH	Alkaline washing	$1.22 \times 10^3$	$4.04 \times 10^2$
Bleaching earth	Degumming	$3.95 \times 10^3$	$1.30 \times 10^3$
Phosphoric acid	Degumming	$2.37 \times 10^2$	$7.82 \times 10^1$

It is consumed  $1.50 \times 10^{-3}$  kg lipase per kg of biodiesel produced, *i.e.* approximately 654 kg of biodiesel are produced per kg of liquid lipase used, which is an acceptable ratio to ensure the economic viability of an industrial plant, according to Novozymes A/S.

Simulation results indicate that the biodiesel produced obeys to the EN 14214 specifications with exception of TAG composition which is slightly above the specification. However, DAG and MAG were not considered in the simulation. Process simulations have several limitations and they do not entirely represent the complexity of a real process. Hence, even though the biodiesel obtained in the simulation might be within the specifications this might not correspond to a real scenario. In a real scenario it would be necessary to perform laboratory tests to be sure that the biodiesel produced in the industrial plant is within the specifications.

### **3.3 Process performance evaluation for PFAD case**

The reaction time is longer in PFAD case - 20 hours vs. 16 hours for CPO. As a consequence, the batch time was increased in 4 hours, being 28 hours in PFAD case. The maximum number of batches per year decreases to 282, causing a reduction in annual biodiesel production.

PFAD is constituted by 85% of FFA and only 12.1 wt% of TAG. This means that the reaction that mainly takes place in the reactor is the esterification of FFA and not the transesterification of TAG. The esterification reaction will form FAME and water as a by-product instead of glycerol. Simulation results show a high production of water, being the glycerol/water stream constituted by 73.2 wt% of water and only 14.8 wt% of glycerol. As 80 wt% of this stream is recycled back to the reactor there is no need to supply water to the reactor apart from the 1<sup>st</sup> batch. However, the water formation also has a negative consequence. Combining the recirculation of 80 wt% of the water/glycerol phase with the recirculation of 50 wt% of biodiesel stream to decrease PFAD melting point, the reactor will

suffer a drastic reduction in its volumetric efficiency.

Consequently, the feedstock flowrate needs to be decreased when the process is running with PFAD. Keeping the same reactors used in CPO case is not possible to run the same flowrate of feedstock per batch. It would be necessary to have more reactors in order to maintain the same feedstock flowrate. This reduction in space-time yield of the reactor causes a significant reduction in biodiesel production per batch, as shown in Table 5.

Regarding the composition of the biodiesel produced, all the criteria from EN 14214 specifications are fulfilled in PFAD case.

In Table 4 is shown the materials consumption for the entire process running with PFAD in "pseudo steady-state".

### **3.4 Comparison between process performance in both cases**

Biodiesel and glycerol production (in a batch and annual basis) for the process running with CPO and running with PFAD can be seen in Table 5.

In Table 5 can be seen that glycerol production is much lower than in CPO case because glycerol is only formed in the transesterification reaction and not in the esterification of FFA, which is the reaction that mainly takes place in PFAD case. The annual glycerol production is 89.6% lower when using PFAD as a feedstock. Another important aspect is that if applying the same conditions in the distillation for glycerol refining as in CPO case, the purity of the glycerol obtained is lower- 70.8% in PFAD case vs. 99.0% in CPO case. This happens because the water/glycerol stream in "pseudo steady-state" has a high content of water- (73.2 wt%), therefore requiring a high energy input in the distillation to evaporate all the water. To achieve a higher glycerol purity in PFAD case, the distillation time could be increased; it could be used a heat transfer agent with a higher heat transfer capacity or the other option is to increase the heat transfer area. But to increase the area, it would be necessary to use another column.

**Table 4 Materials consumption for the process running in "pseudo steady-state" in PFAD case.**

Component	Unit operation	kg/batch	ton/year
PFAD	-	$2.10 \times 10^5$	$5.92 \times 10^4$
Methanol	Transesterification	$2.54 \times 10^4$	$7.17 \times 10^3$
Liquid lipase	Transesterification	$3.08 \times 10^2$	$8.70 \times 10^1$
Water	Alkaline washing	$8.47 \times 10^4$	$1.94 \times 10^4$
	Water washing	$1.03 \times 10^4$	$2.90 \times 10^3$
	TOTAL	$9.50 \times 10^4$	$2.22 \times 10^4$
NaOH	Alkaline washing	$1.73 \times 10^3$	$5.71 \times 10^2$

**Table 5 Biodiesel and glycerol production in a batch and annual basis for CPO and PFAD cases.**

Production		kg/batch	ton/year
Biodiesel	CPO	$3.03 \times 10^5$	$1.00 \times 10^5$
	PFAD	$2.07 \times 10^5$	$5.84 \times 10^4$
Glycerol	CPO	$3.08 \times 10^4$	$1.01 \times 10^4$
	PFAD	$3.73 \times 10^3$	$1.05 \times 10^3$

As a consequence of the reduction in the reactor volumetric efficiency, biodiesel production in a batch basis is 31.7% lower when using PFAD instead of CPO. The reduction in biodiesel production is higher when analyzing the annual production – there is a reduction of 41.7% when using PFAD as feedstock. This can be explained by the longer batch time, which will decrease the maximum number of batches per year.

**Table 6 Energy and power consumption in CPO and PFAD cases.**

Consumption	CPO case	PFAD case
Steam (kg/batch)	$1.35 \times 10^4$	$2.42 \times 10^4$
Cooling water (kg/batch)	$6.35 \times 10^5$	$1.63 \times 10^6$
Chilled water (kg/batch)	$5.55 \times 10^6$	$5.67 \times 10^6$
Power (kWh/batch)	$4.86 \times 10^4$	$6.30 \times 10^4$
Total energy (MJ/batch)	$3.33 \times 10^5$	$4.31 \times 10^5$

In Table 6 can be seen that the overall energy consumption is higher if the process runs with PFAD. Power consumption is approximately 30% higher in PFAD case which is explained

by the longer reaction time for PFAD. However, the difference in total energy consumption is mainly caused by the higher steam and cooling water consumption in PFAD case, since it is necessary a higher energy input in glycerol distillation to evaporate all the water in the glycerol/water stream. Given the low glycerol production and the high energy consumption required to refine it, it should be analyzed if it is economically profitable to keep the glycerol refining section working in PFAD case.

### 3.4 Economics of the process

An estimation of production costs of the process running with each of the feedstocks was made. This estimation is based on prices used in an economical evaluation made for Novozymes A/S on 2010 (with exception of CPO, PFAD and methanol cost price and biodiesel selling price). These prices were not updated to 2013 since the variation was not very significant and given that the aim is not to determine the exact production costs but only to predict the differences in the operation costs when adapting the process to PFAD. The

prices used in the estimation are shown in Table 7.

In Table 8 is shown the cost of the feedstocks, materials and utilities used in both cases and in

Table 9 the revenue obtained from biodiesel and glycerol sales in both cases.

**Table 7 Materials and utilities cost price and products selling price.**

		Price
<b>Materials</b>	CPO	786 \$/MT <sup>[14]</sup>
	PFAD	563.25 \$/MT <sup>[15]</sup>
	Catalyst	50 \$/kg
	Methanol	450 \$/MT <sup>[7]</sup>
<b>Utilities</b>	Steam	12 \$/MT
	Cooling water	0.05 \$/MT
	Chilled water	0.4 \$/MT
	Power	0.1 \$/kWh
<b>Products</b>	Biodiesel	1250 \$/MT <sup>[7]</sup>
	Glycerol	200 \$/MT

**Table 8 Materials and utilities cost per batch for CPO case and PFAD case.**

	Cost per batch (\$)	
	CPO case	PFAD case
<b>Feedstock</b>	$2.48 \times 10^5$	$1.18 \times 10^5$
<b>All materials</b>	$2.91 \times 10^5$	$1.45 \times 10^5$
<b>Utilities</b>	$7.27 \times 10^3$	$8.94 \times 10^3$

**Table 9 Revenue obtained from biodiesel and glycerol sales in CPO case and PFAD case.**

Product	Revenue per batch (\$)	
	CPO case	PFAD case
<b>Biodiesel</b>	$3.79 \times 10^5$	$2.59 \times 10^5$
<b>Glycerol</b>	$6.15 \times 10^3$	$7.46 \times 10^2$
<b>TOTAL</b>	$3.85 \times 10^5$	$2.60 \times 10^5$

In Table 8 can be seen that PFAD is considerably cheaper than CPO - approximately 28.3%. Considering the difference in the price of feedstock and the fact that the batch consumption of all the materials (with exception of water and NaOH) is higher in CPO case, it was concluded that the total materials cost (per batch) is around 50.0% higher in CPO case. In both cases the

feedstock is the biggest contributor for the total materials cost.

On the other hand, utilities cost is around 18.6% higher if the process runs with PFAD since utilities consumption is higher in PFAD case.

Regarding the revenue, given that the biodiesel production is higher in CPO case, the revenue obtained from biodiesel is approximately 31.7% higher than in PFAD case. Since glycerol production is drastically reduced if using PFAD instead of CPO, the difference in the revenue obtained from glycerol in the two cases is high - the revenue obtained from glycerol in CPO case is 87.9% higher than in PFAD case. Overall, the total revenue of the process is 32.6% lower if using PFAD.

Given that feedstock cost corresponds to 70-95% of total operation costs' it can be considered that operation costs are approximately the same as the materials cost. Therefore, it can be concluded that when the industrial plant runs with PFAD the operating costs are lower but the revenue is also lower. On the other hand, operating exclusively with CPO has a batch operating cost 50% higher than in PFAD case but the revenues are only around 33% higher per batch. The optimal operating point should be somewhere between using exclusively PFAD and using only CPO. An economical study should be done to determine the optimum number of batches the industrial plant should run with each feedstock, in order to obtain a balance between operating costs and revenue.

## 4 Conclusions

The combination of crude palm oil and PFAD in a flexible enzymatic process was evaluated and it has been accomplished but with significant differences in the process throughput. When using PFAD instead of CPO there is a drastic reduction in the volumetric efficiency of the reactor which causes a reduction in biodiesel production and overall process productivity.

Having an enzyme that could maintain its activity at temperatures higher than 39 °C would improve the productivity of this flexible process. Hence, collaboration between process engineering and research in genetic engineering and protein modification is vital for the optimization of this process.

## 5 References

- [1] P. M. Nielsen, J. Brask, and L. Fjerbaek, "Enzymatic biodiesel production: Technical and economical considerations," *European Journal of Lipid Science and Technology*, vol. 110, no. 8, pp. 692–700, Aug. 2008.
- [2] L. Fjerbaek, K. V Christensen, and B. Norddahl, "A review of the current state of biodiesel production using enzymatic transesterification.," *Biotechnology and bioengineering*, vol. 102, no. 5, pp. 1298–315, Apr. 2009.
- [3] G. Knothe, J. Van Gerpen, and J. Krahl, *The Biodiesel Handbook*. AOCS Press, 2005.
- [4] S. Al-zuhair, "Production of biodiesel : possibilities and challenges," *Biofuels, Bioproducts & Biorefining*, no. April, pp. 57–66, 2007.
- [5] M. J. Haas, A. J. McAloon, W. C. Yee, and T. a Foglia, "A process model to estimate biodiesel production costs.," *Bioresource technology*, vol. 97, no. 4, pp. 671–8, Mar. 2006.
- [6] Y. Xu, "Process Technology for Immobilized Lipase- catalyzed Reactions," Technical University of Denmark, 2012.
- [7] P. M. Nielsen, S. Träff, and H. C. Holm, "Enzymatic catalyzed FAME now more cost effective than chemical catalyst," *AOCS 103rd Annual Meeting*, 2012.
- [8] J. Brask, M. L. Damstrup, P. M. Nielsen, H. C. Holm, J. Maes, and W. De Greyt, "Combining enzymatic esterification with conventional alkaline transesterification in an integrated biodiesel process.," *Applied biochemistry and biotechnology*, vol. 163, no. 7, pp. 918–27, Apr. 2011.
- [9] "Vegetable oils melting points," *Engineering Toolbox*. [Online]. Available: [http://www.engineeringtoolbox.com/oil-melting-points-d\\_1088.html](http://www.engineeringtoolbox.com/oil-melting-points-d_1088.html). [Accessed: 22-Jan-2013].
- [10] S. H. Goh, Y. M. Choo, and S. H. Ong, "Minor Constituents of Palm Oil," *Journal of the American Oil Chemists' Society*, vol. 62, no. 2, pp. 237–240, 1985.
- [11] E. Crabbe, C. Nolasco-Hipolito, G. Kobayashi, K. Sonomoto, and A. Ishizaki, "Biodiesel production from crude palm oil and evaluation of butanol extraction and fuel properties," *Process Biochemistry*, vol. 37, no. 1, pp. 65–71, Sep. 2001.
- [12] R. binti M. Zin, "Process Design in Degumming and Bleaching of Palm Oil," Universiti Teknologi Malaysia, 2006.
- [13] G. Güzel, "Sustainable Biocatalytic Biodiesel Production- A Thermodynamic Analysis," Arhus University, 2012.
- [14] "Crude palm oil price," *Palm Oil HQ Lti Ltd*, 2009. [Online]. Available: <http://www.palmoilhq.com/palm-oil-prices/>. [Accessed: 18-Jun-2013].
- [15] "PFAD price," *Commodity 3*, 2013. [Online]. Available: <http://www.commodity3.com/instrument/PFA0MYM1/pfad-palm-fatty-acid-distillate>. [Accessed: 18-Jun-2013].