Mill Integrated Conversion of Palm Biomass to Commodity Intermediates

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Mill Integrated Conversion of Palm Biomass to Commodity Intermediates

PDEng Design Project

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Keywords: second-generation chemicals, biorefinery, fermentable sugars, palm oil, solid biomass.

Table of contents

Summary

The current production of palm oil produces large amounts of biomass residues, namely empty fruit bunches, shells, fibres, fronds, and trunks. The current worldwide demand for biomass-derived biochemical intermediates, provides the opportunity for converting these biomass residues into higher value products. In this design project, the feasibility of 2nd generation intermediates obtained from oil palm residues has been evaluated. The focus of the project has been the integration of a conversion process into the palm oil milling facilities, considering availability of residues at different scales. Thus, logistics implications regarding biomass sources, which are considered to have a large impact in biomass cost, are taken into account in the evaluation.

After a preliminary evaluation of different alternatives, a fermentable sugars product is found as the most promising product for a 2nd generation palm-based biorefinery, based on economic margin, market opportunity, market size, process safety, and environmental impact. As part of the process design for the biomass transformation, it was found that a pre-treatment step is necessary due to the recalcitrant nature of the material. Several pre-treatment alternatives and their process implications were evaluated, indicating that acid catalysed steam explosion is the most suitable alternative. The resulting cellulosic fraction undergoes enzymatic hydrolysis to give fermentable sugars. Additional sugars are recovered when biomass available form plantation operations is available.

A preliminary evaluation of the process indicates that the benefits related to economies of scale are overwhelmed by the biomass transportation costs that larger scales imply, finding an optimum scale of three mills (biomass derived from operations of three palm oil mills) with capacity of 60 tonne FFB/h each. Following these initial findings, a more detailed computer model of the process was built in order to compare cases with biomass derived from three and ten mills.

An economic evaluation indicates that the process is profitable in both cases, with a payback in the third year of operation. As in the preliminary estimations, the proposed design results in total processing costs lower for the smaller scale, with 1 074 and 1 120 RM/tonne of sugar for the three- and ten-mill cases respectively. Sensitivity analyses indicate that the process profitability is more robust to changes on transportation distance amongst mills and sugar price when processing biomass derived from three mills.

Furthermore, after a Life Cycle Assesment (LCA) it is found that the environmental impact of palm oil, as in EDIP 2003 methodology, is greatly reduced when the proposed process is included. The degree in which the impact is decreased depends on the considerations taken in the assessment, Bearing in mind the fermentable sugar product is a sugar replacement, the environmental impact of palm oil can be reduced at least 39%. Although the assessment is based on rough considerations, given the scope of this design project this impact reduction is considered a good initial estimation of the benefits it can bring to the palm oil production.

Overall, it is found that the proposed process can bring an economic benefit to palm oil mills, increasing the availability of renewable feedstocks for the chemical industry in Malaysia. It is found that larger scales result in higher sugar production costs when transportation of biomass is considered. Thus, the profitability of the proposed process at lower scales is more robust to changes in fermentable sugar product price and transportation distance. Thus, it is considered that the process at a three-mill scale is a better alternative.

List of Abbreviations

AD	Anaerobic Digestion
AFEX	Ammonia Fibre Explosion
AISE	Acid Impregnated Steam Explosion
Ak	Alkaline Treatment
CAPEX	Capital Expenditure
C_{BM}	Biomass Cost
Cbuildings	Buildings Cost
Ccap	Capital Charge
CGEE	Brazilian Centro de Gestao e Estudos Estrategicos
C_{gral}	General Expenses
C_{lab}	Total Labour Cost
C_{maint}	Maintenance Cost
COD	Chemical Oxygen Demand
$C_{offsite}$	Offsite Facilities Cost
C_{overh}	Overhead expenses
CPKO	Crude Palm Kernel Oil
CPO	Crude Palm Oil
C_{RM}	Raw Material Cost
C_{site}	Site Development Cost
C_{tax}	Insurance and Taxes
CTBM	Total Bare Module Cost
C_{TDC}	Depreciable capital
C_{waste}	Waste Disposal Cost
C_{wc}	Working Capital
DA	Dilute Acid
DPC	Direct Production Cost
EFB	Empty Fruit Bunches
FFB	Fresh Fruit Bunches
FV	Future Value
HMF	Hydroxy-Methyl Furfural
i	Interest Rate
IRR	Internal Rate of Return
LCA	Life-Cycle Assessment
LHW	Liquid Hot Water

LP	Low Pressure
LTC	Long Term Contract
MICCI	Mill Integrated Conversion of palm biomass to Commodity Intermediates
MP	Medium Pressure
NBS	National Biomass Strategy
NPV	Net Present Value
NREL	U.S. National Renewable Energy Laboratory
NRTL	Non-Random Two-Liquid Thermodynamic Model
OPBC	Oil Palm Biomass Centre
OPF	Oil Palm Fronds
OPT	Oil Palm Trunks
POME	Palm Oil Mill Effluent
PV	Present Value
SE	Steam Explosion
TCI	Total Capital Investment
TPC	Total Production Cost
TS	Total Solids (Soluble + Insoluble)

1. Description of the design

1.1. Background

Malaysia is one of the largest palm oil producers in the world, reaching a production of 19 million tonnes of crude palm oil in 2011 (Malaysian Palm Oil Board, 2012). As a result, large amounts of biomass residues are constantly generated at palm oil mills. The Malaysian government, through the National Biomass Strategy (NBS), puts into focus the opportunities inherent in the oil palm biomass residues and the foundations for implementing their conversion into higher value products (Agensi Inovasi Malaysia, 2011). Parallel to the NBS publication, the Malaysian Oil Palm Biomass centre (OPBC) has been created out of a public and private partnership, with the objective to develop a more sustainable palm-based industrial sector for the production of bio-derived chemicals, fuels, and energy. Both academia and industry have embarked in this joint venture to make Malaysia's oil palm sector an economic and sustainable world leader (BE-Basic, 2011).

The OPBC joint venture has organized its R&D under five technology programs related to the oil palm biomass value chain:

1. Sustainable plantation and harvest
2. Logistics and biorefining
3. Conversion to biochemicals, materials, biofuels, and bioenergy
4. Nutrient recycling, soil and water quality
5. Energy and carbon dioxide integration

As part of OPBC, this exploratory design project is mainly related its 3rd programme on conversion to biochemicals, materials, biofuels and bioenergy.

Imbedded in the mill context and considering the biomass residues available from its operations, it is the goal of this conceptual design project to investigate the techno-economic feasibility of 2nd generation intermediates for further processing to chemicals, fuels, and energy for practical application. These targeted intermediates are to be sold to a third party for further processing, whereas ligneous residues will be utilised for the energy demands of the mill.

In order to achieve the goal of the design project, an evaluation of various pretreatment/hydrolysis technologies as well as a preliminary model-based analysis towards pilot plant scale will be performed. As it is common that mills from the same crude palm oil producer are located in nearby regions, the scaling of the design is a possible scenario. Therefore, as part of this design project, the trade-off between the economy of scale and transportation cost will be evaluated. Furthermore, the raw material for the 2nd generation intermediates production can be obtained from mill derived biomass (empty fruit bunches, shells, fibres) and/or plantation derived fronds, and trunks. In order to compare all these options this design project will initially compare four different cases: (1) pretreatment at the mill with mill derived biomass, (2) pretreatment at the mill with plantation plus mill derived biomass, (3) pretreatment at the mill with biomass derived from multiple mills, and (4) pretreatment at the mill with biomass derived from multiple mills and the plantation related to them. The result of this evaluation will be used to make a selection and proposal, delivered as Basis of Design. Thereafter, the process is designed with more detail under the chosen scenario(s), and evaluated based on its economic implications.

Furthermore, as that the palm oil industry has been target of numerous critiques regarding environmental implications of intensive oil production, it is important to know the environmental impact the new design can have. Hence, this project will be complemented with a life cycle analysis (LCA) considering the designed process integrated to milling operations. The results of this analysis are to be compared to those derived from typical palm oil milling operations to have a preliminary estimation of the effect it can bring to the palm oil industry.

1.2. Analysis of the problem

Oil palm *Elaeis guineensis*, indigenous to west Africa, is a palm largely utilized for oil extraction from its fruits. The fruits are contained in what is called fresh fruit bunches (FFB) that commonly obtain weights of 10 to 40 kg. The fruits in these bunches are composed of a seed (kernel) covered by an oily pulp. Oil can be extracted from both of these parts, namely crude palm oil (CPO) from the pulp and crude palm kernel oil (CPKO) from the fruit's seed (Shuit *et al.*, 2009).

At palm oil mills, FFBs are processed for the extraction of oil and consequently, empty fruit bunches (EFB) are obtained as a by-product, accounting up to 23% (w/w) of FFB. Shells, fibres, and kernels are also produced after the fruit processing, although in smaller amount when compared to EFB, with 5.5%, 13.5% and 6% (w/w) of FFB respectively (Malaysian Department of Environment, 1999). Additionally, other biomass by-products, such as oil palm fronds (OPF) and oil palm trunks (OPT), are constantly generated at the plantation sites (Agensi Inovasi Malaysia, 2011).

Traditionally, EFB were incinerated and the derived ashes were used as fertilizer. This practice has decreased since EFB demand for mulching purposes has increased to reduce the fertilisation cost on the plantations. Additionally, EFB incineration has been strongly discouraged by the Malaysian Department of Environment, due to the related emissions (Er *et al.*, 2011). The fibre and shell residues are commonly used for steam and electricity generation for in-house use, which makes mills energy self-sufficient. Yet, other uses for fibres and shells have been suggested, such as composite materials for furniture and automobile components as well as hardening material for roads (Er *et al.*, 2011). On the other hand, the fronds are commonly left as topsoil replacement or natural fertilizer. The trunks, becoming available after a production lifetime of *ca.* 25 years, are also left on the field as fertilizer. An exception to this practice is the use of trunks as raw material for the wood industry (Agensi Inovasi Malaysia, 2011).

Besides the mentioned uses of palm biomass solid residues, there are several research projects focusing on the use of these materials, particularly EFB. Among biological treatments of the waste, the production of sugars for ethanol (Shamsudin *et al.*, 2012) or polylactate production (Hassan *et al.*, 2009) and cellulases (Alam *et al.*, 2009) have been described in the literature. Catalytic processes for liquid and gaseous biofuels have also been proposed (Chew *et al.*, 2008).

Considering the materials composition, the process for the biomass residues transformation implies the breakdown of the lignocellulosic components. If plantation biomass is also considered the extraction of a sugar-rich sap is also needed. The steps required for biomass conversion are simply schematized in a block diagram that is shown in Figure 1. The concept considers that the un-converted lignin component will be used for the energy demands of the mill in order to keep its current energy self-sufficient nature.

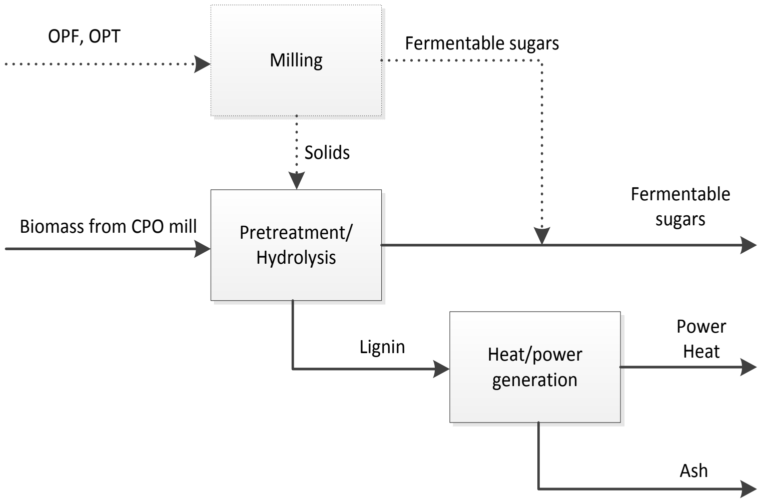


Figure 1. Block diagram of necessary steps for biomass transformation into products
Dotted lines apply for biomass derived from plantations. OPT: Oil palm trunk, OPF: Oil palm fronds, EFB: Empty fruit bunches, CPO: Crude palm oil.

Mill and plantation derived biomass residues are mainly lignocellulosic material that is highly recalcitrant in its natural state. This characteristic of the biomass makes it hard to recover the carbohydrates that form part of its structure. Hence, in this design project special attention will be given to the required pretreatment and hydrolysis operations. Several pretreatment methods have been developed in the last decades. Their implementation at large scale has been limited due to high energy costs and sugar degradation into products that can be inhibitors to microbial or enzymatic activity. The selection of an appropriate approach for the process will be presented in Section 2.7.

This design project will then initially evaluate four cases and compare these to a typical mill or Base Case. The process implications will be evaluated and compared for each case, which differ on the type of input biomass, the scale of biomass sources and the type of POME treatment. Complementary to Base Case, anaerobic POME treatment with biogas recovery is included as Base Case Plus. This consideration is based on the fact that biogas recovery is already established as part of future palm oil mills, although not commonly available in current mills. See Table 1 as well as Figures 2 through 4 for an overview of the different cases. As a consequence of using biomass from different sources and locations, the logistics and thus transportation costs are different, and these will also be considered in this study (see Figure 5).

Table 1. Sources of residual biomass , scale of intermediate production, and POME treatment for all the different cases

Case	Type of biomass	Scale	POME treatment
Base Case	Shells and fibres	Single mill	Open air lagoon
Base Case Plus	Shells and fibres	Single mill	Anaerobic Digestion
Case A	Shells, fibres, and EFB	Single mill	Anaerobic Digestion
Case B	Shells, fibres, EFB, OPT and OPF	Single mill	Anaerobic Digestion
Case C	Shells, fibres, and EFB	Multiple mills	Anaerobic Digestion
Case D	Shells, fibres, EFB, OPT and OPF	Multiple mills	Anaerobic Digestion

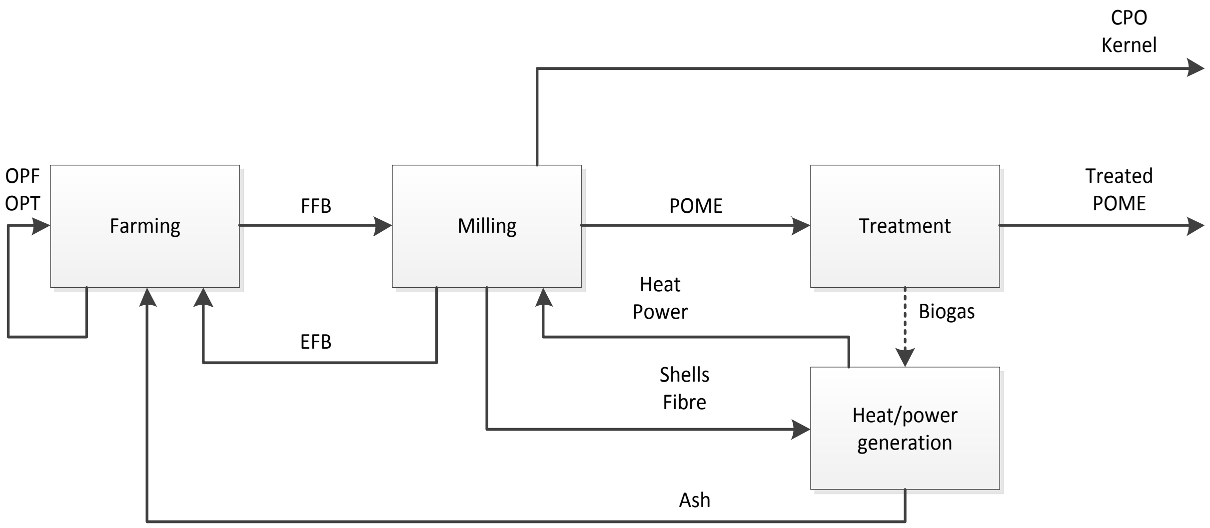


Figure 2. Block diagram for Base Case and Base Case Plus

Dotted lines apply for Base Case Plus. OPT: Oil palm trunk, OPF: Oil palm fronds, FFB: Fresh fruit bunches, EFB: Empty fruit bunches, CPO: Crude palm kernel oil, POME: Palm oil mill effluent.

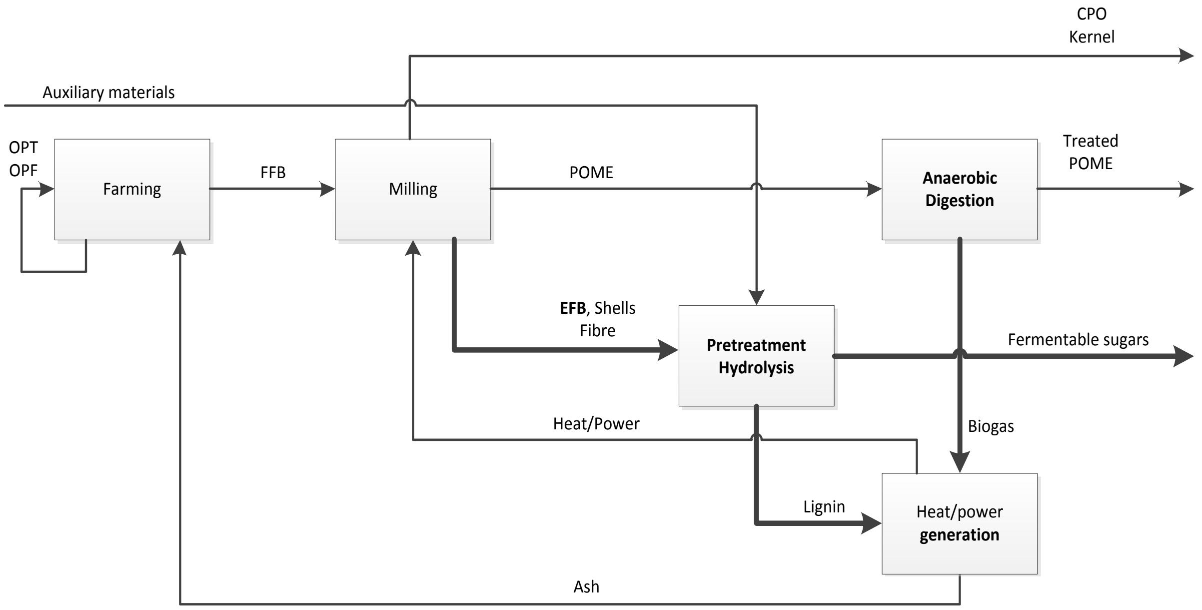


Figure 3. Block diagram for Cases A and Case C (multiple mill scale)

OPT: Oil palm trunk, OPF: Oil palm fronds, FFB: Fresh fruit bunches, EFB: Empty fruit bunches, CPO: Crude palm kernel oil, POME: Palm oil mill effluent.

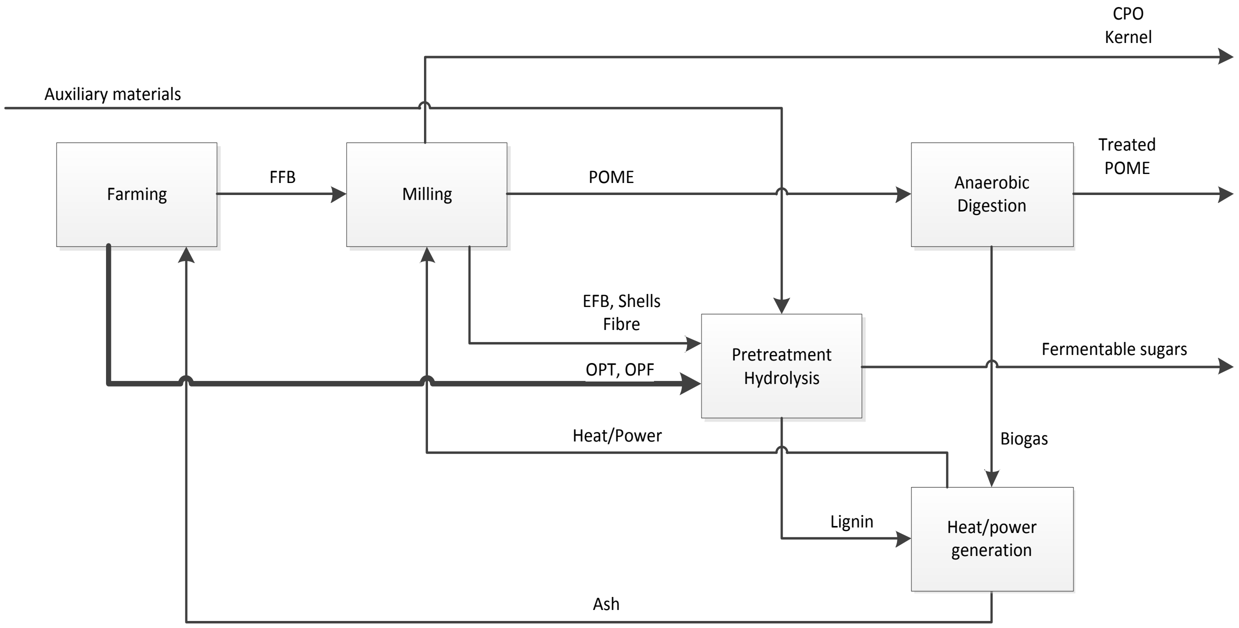


Figure 4. Block diagram for Case B and Case D (multiple mill scale)
OPT: Oil palm trunk, OPF: Oil palm fronds, FFB: Fresh fruit bunches, EFB: Empty fruit bunches, CPO: Crude palm kernel oil, POME: Palm oil mill effluent.

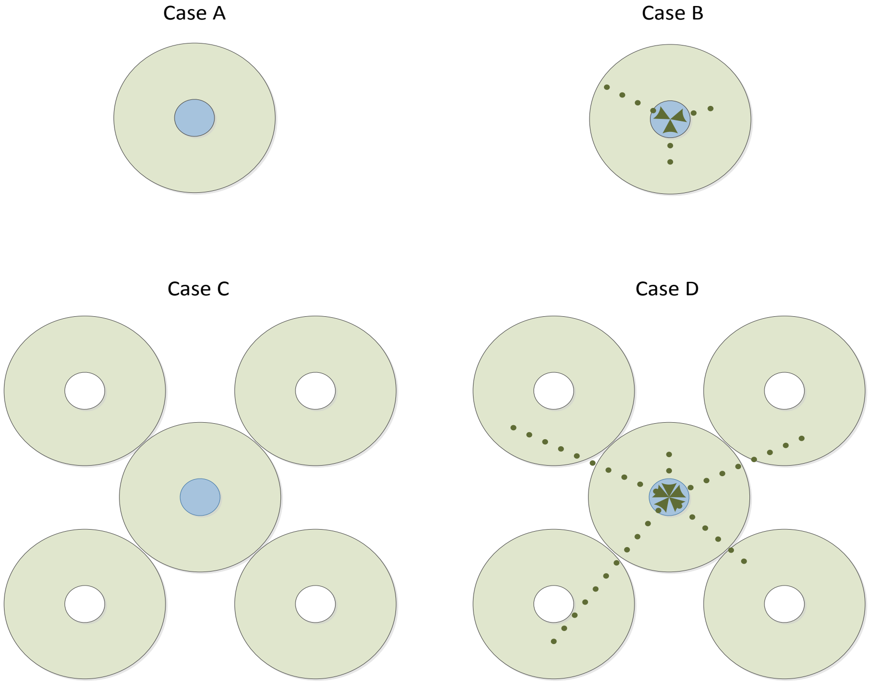
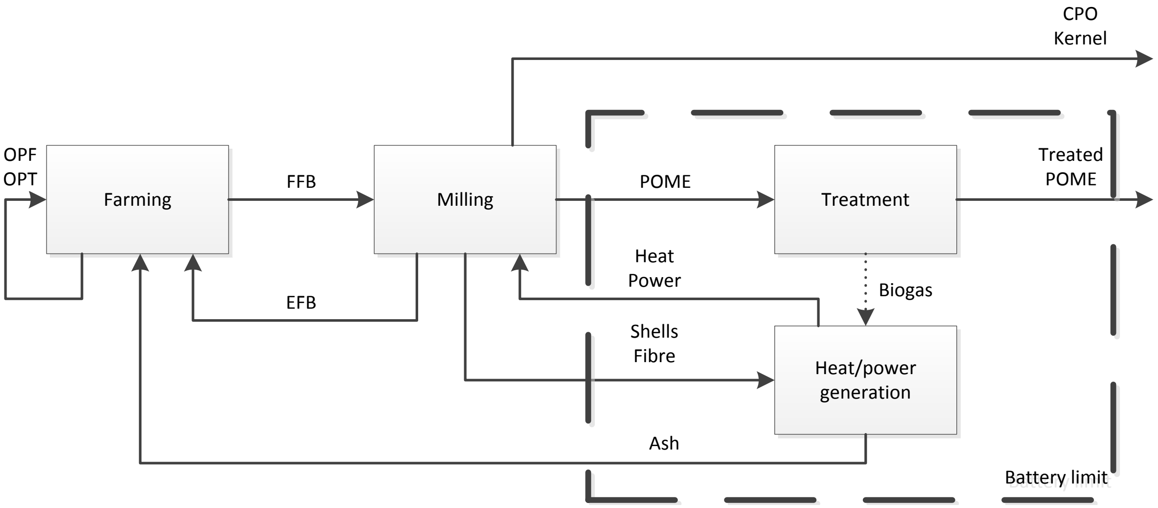


Figure 5. Transportation logistics for the for cases
Blue circles: Intermediates producing mill, white circle: typical mill, dotted line: oil palm frond and oil palm trunk transportation.

1.3. Battery limits

Palm oil operations related to farming and oil extraction (FFB sterilization and threshing, oil extraction and clarification, and kernel recovery) will not be altered and are outside of the scope of this design project, so they are placed outside of the battery limits. The biomass residue materials are thus considered as battery limit crossing streams, Figure 6, with proper cost allocation as described in Section 8.1Raw Materials Cost.



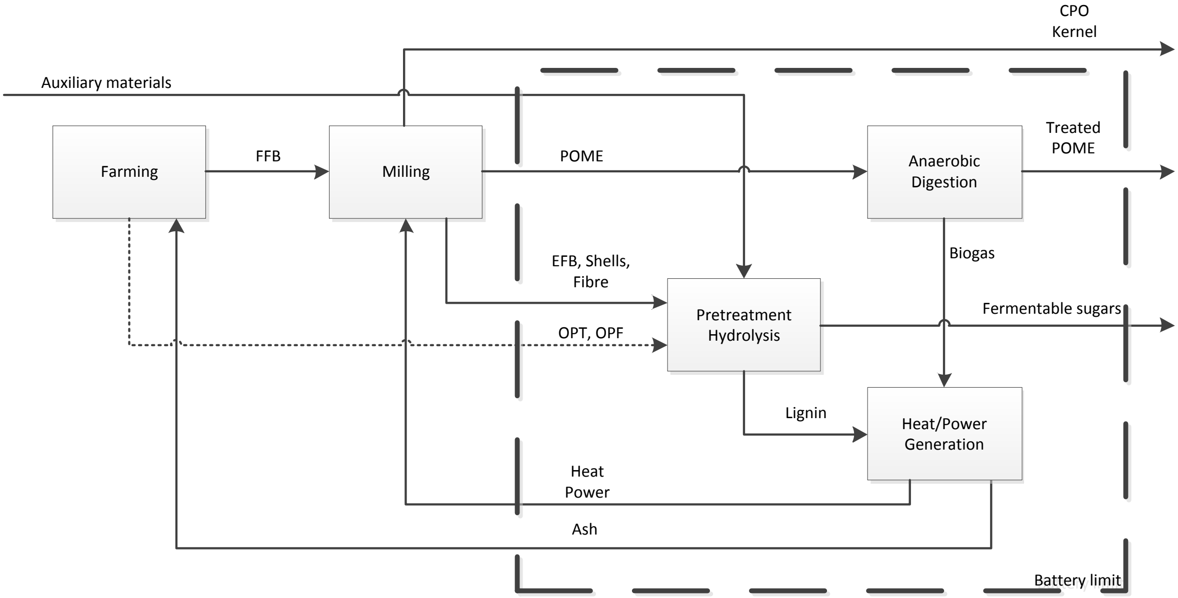


Figure 6. Block diagrams with battery limits

Top: Base Case (dotted line applies for Base Case Plus), Bottom: Cases A, B C and D (dotted line applies for Cases B and D). OPT: Oil palm trunk, OPF: Oil palm fronds, FFB: Fresh fruit bunches, EFB: Empty fruit bunches, CPO: Crude palm kernel oil, POME: Palm oil mill effluent.

Steam and electricity generation is considered inside the battery limits in all cases, whereas the rest of the utilities are not. To make a fair comparison, a cost is allocated to any other of the utilities. Additionally, since biomass residue materials are assumed to have a cost, the amounts of steam and electricity derived from them, although being intended for use inside the mill, are considered products of the design and a price is assigned to them (see Section 8.1.2 for details). Any water treatment on fresh river water is considered outside of the battery limits as it is considered an operation within the palm oil milling. Transportation of biomass from plantation into the biorefinery is considered for OPT, OPF. Transportation amongst mills is considered for biomass available at the mills.

2. Process Definition

After an evaluation of several product alternatives for the palm oil mill (Appendix 1 – Product Alternatives), it is decided that mill integrated conversion of palm biomass to commodity intermediates (MICCI) process has fermentable sugars as target product. The conversion of this lignocellulosic material into sugars requires the breakdown of its polymeric components, mainly hydrolysis of cellulose and hemicellulose into monomeric sugars. Cellulose hydrolysis alternatives are presented in section 2.1. However, as lignocellulosic material is highly recalcitrant in its natural state, it is usually subjected to pretreatment step(s) to improve its further conversion into fermentable sugars. Several alternatives have been developed for pretreatment of lignocellulosic material; the main alternatives are presented and an appropriate method is selected in section 2.2. As most pre-treatment methods result in the formation of toxic compounds for hydrolysis, conditioning methods will be evaluated in section 2.3. After the selection of the main process sections, different process alternatives based on biomass logistics are evaluated. In section 2.7 a process alternative is selected.

2.1. Cellulose hydrolysis

There are typically two routes for hydrolysis of cellulose, a chemical and an enzymatic route. Chemical hydrolysis is a simple process that requires the use of concentrated acid catalyst for the breakage of the glucosidic bond. As a consequence of the severity of this treatment, product degradation often takes place. Product degradation not only implies the loss of valuable sugars, but also the formation of compounds that are not desirable in a product stream intended for industrial fermentation (degradation products can be, for instance, growth inhibitors of ethanol producing microorganisms). On the other hand, enzymatic hydrolysis is a specific method for breaking the glucosidic bond holding the cellulose structure, and thus no by products are formed. Its specificity, the mild conditions required for the reaction and the rapid improvements achieved in the industrial enzyme field have made it the method of choice for cellulose hydrolysis.

For the enzymatic hydrolysis of cellulose two main options can be considered, (1) in-house production of the enzymes from a side-stream of the pretreated biomass and (2) use of commercially available enzymes. In-house production of enzymes is typically considered an option for large scales, where an increase in the capital expenditure might be overcome by lower operational costs. However, it has been reported in the literature that enzyme costs delivered by commercial suppliers has been decreasing with the occurrence of partnerships amongst enzyme producing companies, research institutes (like NREL), and ethanol producers (like POET Ltd). As result, early in 2012, the release of the latest Novozyme cellulosic enzyme was released with 80% price drops in comparison with the previous two years (Leber, 2012).

Based on current status of commercial enzyme preparations, acquiring enzymes from commercial partners is considered a better option for implementation at a typical mill. Several enzyme preparations are available commercially from companies like Novozymes, DSM and Genencor.

2.2. Pretreatment methods

Different pretreatment methods and their combinations have been developed, each with different advantages and disadvantages. In this design project, high importance is given to the production of monomeric sugars from both pentoses and hexoses with high yields. Furthermore, the project requires that the pretreatment method is a simple process that can be easily adapted into the palm oil mill setting. These required pretreatment characteristics are translated into selection criteria, as presented in Table 2.

In order to improve enzymatic digestibility of cellulose, the strategy to follow can be to first solubilize lignin or hemicellulose. Alternatively, physical treatment of the lignocellulosic material can be used to open up the polymeric structure. In Appendix 2 – Pretreatment Alternatives, a brief description of the main pretreatment technologies used for facilitating cellulose hydrolysis can be found.

Based on the advantages and disadvantages of the different pretreatments found in the literature, some alternatives have been directly eliminated based on extensive energy requirements and low reaction rates. Steam explosion (SE), acid impregnated steam explosion (AISE), ammonia fibre explosion (AFEX), liquid hot water (LHW), dilute acid (DA) and alkali (Ak) treatments have been further considered in this chapter.

The sugar yields reached with any pretreatment is a key characteristic of the method. Considering this, a literature study was performed on reported yields obtained with palm biomass. It was found that pretreatment of oil palm biomass has been mostly focused on EFB. However, the extent to which different pretreatment technologies have been studied on this biomass residue is low compared to other biomass types like corn stover or switchgrass. As a result, yields obtained with EFB are generally low when compared to those obtained with other biomass types under optimized conditions. The reached yields can be expected to grow when more experience is gained.

Recently, a comparative study of different pretreatment technologies, focusing on conversion yields and sugar degradation reactions was published by Kumar *et al.* (2011). Yields presented in the mentioned comparative study (Kumar *et al.*, 2011), which were obtained experimentally by other authors or used by Kumar for simulation, are presented in Table 2 along with EFB yields from different pretreatments found in the literature. A list of reviewed literature is presented in Appendix 2 – Pretreatment Alternatives.

Table 2. Comparison of yields obtained with different pretreatment technologies

Pretreatment	EFB			Other biomass*		
	Yield^{+}	Temp. (°C)	Source	Yield^{+}	Temp. (°C)	Source
SE	0.31	140	a	0.82	180	g
AISE	0.82^{++}	140	a, b	0.88	190	i
DA	0.69	120	c, d	0.92	180	g
Ak	0.72	128	e	0.83	180	g
AFEX	0.85	135	f	0.95	90	h
LHW	-		-	0.87	180	g

a: Shamsudin *et al.* (2012), b: (Zimbardi *et al.*, 2007), c: (Rahman *et al.*, 2007), d: (Razali *et al.*, 2009), e: (Han *et al.*, 2011), f: (Lau *et al.*, 2010), g: (Kumar *et al.*, 2011), h: (Laser *et al.*, 2009), i: (Carrasco *et al.*, 2010)

$^{+}$Yields are % of theoretical mass conversion yields, obtained after pretreatment and enzymatic hydrolysis. These are adapted from separate xylose and glucose yields, considering xylan and glucan content in EFB.

$^{++}$ The AISE yield for EFB is derived from the increment in SE yield using acid impregnation, as reported by (Zimbardi *et al.*, 2007). The increment is applied to the sugar yield form SE.

* Cornstover, switchgrass and grass straw.

The yield discrepancy amongst pretreatment of EFB and other biomass types can be intrinsic of the treatment severity, of the biomass presentation (particle size and moisture) or of the biomass *per se* (composition and structure). However, pretreatment methods can deliver similar performance that can be generalized for certain biomass groups (e.g., acid catalysed steam explosion alone is effective for agricultural residues, whereas an extra delignification step is often needed for softwoods) or even for all groups (e.g. for catalysed treatments, high moisture contents are favourable for all biomass types since it allows a better catalyst distribution in the biomass bulk). Overall, biomass digestibility does not only relate to fibre composition, but also to porosity, particle size and cellulose crystallinity (Park *et al.*, 2010). The yields achieved with similar biomass types, like agricultural residues, could thus be considered as attainable yields for EFB and other oil palm biomass.

2.2.1. *Selection of pretreatment*

Yields are a good point of comparison for selecting a pretreatment strategy. However, as each alternative has different process implications, yields cannot be the sole factor affecting pretreatment selection. Thus, each alternative is compared to each other based on selection criteria, as presented in Table 3.

Table 3. Comparison of pretreatment alternatives based on selection criteria

Criterion	Alternative					
	SE	AISE	DA	Ak	AFEX	LHW
High sugars yield	-	**+**	+	0	+	0
Complexity	+	**0**	0	0	-	0
Conditioning requirement	0	**0**	-	0	+	-
Low energy consumption	+	**+**	0	0	-	0
Low extra material costs	+	**0**	0	-	+	+
Low safety and environmental risk	+	**0**	0	0	-	+
Simple lignin recovery	+	**+**	+	-	0	+

Positive (+), negative (-) or neutral (-) compared to each other.

Steam explosion (SE), acid impregnated steam explosion (AISE), ammonia fibre explosion (AFEX), liquid hot water (LHW), dilute acid (DA) and alkali (Ak) treatments

Steam explosion alternatives are the most appealing, with DA, AFEX and LHW scoring in mid-level. Alkali treatment scores lowest, although it has been widely studied, mainly for pulp and paper production. In alkali treatment a fraction of lignin is solubilized in the liquid stream so an extra precipitation step for its recovery and use for energy generation would be required. In both alkali and acid treatments, which operate in relatively dilute conditions, the use of larger amounts of water implies more energy for heating up to the elevated operating temperatures, besides the increase in operating volumes and thus capital investment.

Inhibitors production or the need for conditioning is relevant considering the subsequent step to pretreatment. Enzymatic hydrolysis is susceptible to product inhibition, and has an optimum pH in the range of 5. Also, as the product sugar stream is intended for industrial fermentation, no microbial activity inhibitors are desirable. Commonly, the fact that no conditioning is required after pretreatment is one of the main advantages referred to AFEX. However, there are other implications in AFEX pretreatment. If a washing step takes place after AFEX pretreatment, a fraction of the lignin and solubilized xylose oligomers are lost in the washing stream and would need to be recovered, especially if lignin is intended for energy generation. On the other hand, when no washing step is performed, glucose yields are lower (compared to yields achieved with washing or other pretreatments), which is thought to be due to the presence of xylose oligomers liberated during AFEX pretreatment, and which are inhibitory to cellulases (Garlock *et al.*, 2011).

LHW also has similar implications regarding inhibition of enzymatic hydrolysis. When the fibres are not washed, sugar yields are much lower. If washing is used, any inhibitory effect on cellulases is avoided. However, higher levels of furfural can be present in the pretreatment liquor, indicating xylose degradation and also implying possible need of more specialized conditioning. Furthermore, LHW has been regarded as not feasible at large scale and is rather thought of as technology available only in a long-term.

During steam explosion some acetic acid might be liberated and thus the pH is decreased. However, the impact of the formed acid is considered low when compared to conditioning needed with acid and alkali treatments that result in higher phenolics or furan release. Even in the case of acid impregnated steam explosion, furan release has been found at levels which are considered not inhibitory (Carrasco *et al.*, 2010), thus conditioning besides pH increase is not necessary. However, some cellulose inhibition was found with unwashed fibres after AISE, thus a washing step is preferred to remove possible xylose monomers and oligomers that could be inhibitory for cellulases.

Process complexity is mainly referred to the ease of process integration with the already existing mill facilities. Because steam is already available at the mill, and no other catalyst is necessary in this treatment, it is clear that steam pretreatment scores highest in this criterion. However, the lower yields achieved with this pretreatment make it necessary to turn to AISE treatment. It has been mentioned in the literature that AISE is closest to commercialization when compared to other pretreatments, with pilot scales plants in countries like Canada, France, and Sweden (Hahn-Hägerdal *et al.*, 2006). AFEX on the other hand is considered the most complex of the alternatives, since it requires catalyst recovery to be economically feasible. Ammonia recovery, even with the relatively new quenching approach, is still energy intensive and adds up complexity to the process.

SE is considered the alternative that adds less safety and environmental hazards. Regarding operating pressure, it is considered that since the mill already operates with high pressures, SE implies no additional major risk. Although, SE scores higher overall (Table 3), it is decided not to work with this pretreatment without catalyst addition as the considerably low sugar yield is a show-stopper with high economic impact. As a result, acid impregnated steam explosion pretreatment will be further considered.

2.3. Conditioning

Additionally to sugars, other compounds are released during pretreatment. The type and extent of released compounds depends on the type and severity of the pretreatment. At severe conditions, furfural and hydroxymethyl-furfural (HMF) are obtained from the degradation of pentoses and hexoses respectively, while phenolic compounds are the main result of lignin degradation. Additionally, acetic acid might also be released during pretreatment, particularly during acidic treatments.

As that the main objective of the pretreatment step is the increase of enzymatic digestibility of the biomass, it is clear that any compound that has a detrimental effect on it, is considered an inhibitor. Furthermore, as the target sugar product is intended for industrial fermentations, it is necessary to remove or reduce the presence of microbial activity inhibitors in the end product.

Taking microbial inhibition studies found in the literature, furfural, HMF and phenolic compounds, mainly of 4-hydroxy benzoic acid and vanillin derived from lignin degradation, are considered inhibitors (Pienkos *et al.*, 2009). It has been reported however that concentrations up to 2 g/L of 4-hydroxy benzoic acid and 1g/L of vanillic acid are not inhibitory for either microbial growth or ethanol production (Palmqvist *et al.*, 2000). Acetic acid has a special role due to its inhibitory effect mainly related to pH. At lower pH, acetic acid can penetrate microbial membranes and de-regulate the proton balance, thus pH must be neutralized to avoid inhibition related to acetic acid.

There are three main types of conditioning or detoxification:

1. *Biological conditioning* refers to the use of microorganisms or their parts to remove inhibitory compounds. Enzymes like laccasas and lignin peroxidases, or microorganisms producing them, to remove monoaromatic phenolics have been reported. Also, furfural and HMF have been removed with fungal strains, like *Coniochaeta ligniaria.* Other microorganisms, like *Ureibacillus thermosphaericus* has been found to oxidize furfural and HMF to less toxic compounds (Pienkos *et al.*, 2009).

2. *Chemical conditioning* is generally done by neutralization of the acids or alkali used for pretreatment, to meet the appropriate pH for enzymatic hydrolysis and/or fermentation. In the case of acid neutralization, the most common method is overliming, in which the pH is raised to 10 and lowered to 5.5. It is common to use $Ca(OH)_2$ as it is cheap, although it results in gypsum production and some sugar losses with gypsum removal (Pienkos *et al.*, 2009). As an alternative, the use of NH_4OH has been used in the design by Humbird *et al.* (2011), where the pH is directly raised from pretreatment pH to 5, however the economic implications have not been thoroughly studied.

3. *Physical conditioning* involves the physical removal of the toxic compound. Different methods have been considered, such as liquid extraction, electrodyalisis, supercritical fluid extraction, ion exchange and even solid phase extraction (Pienkos *et al.*, 2009).

Conditioning with techniques such as anion exchange has been proved successful and effective for processes with concentrated acid, in which the recovery of the catalyst is imperative. However, its implementation for processes with diluted catalyst, where catalyst recovery is often considered not cost-effective, has been less supported (Hamelinck *et al.*, 2003). In this later case, it is common to turn to just neutralization, most commonly with lime.

Additionally, it is relevant to remark that some of the mentioned inhibitors are volatile. Under this consideration, designs with flash cooling of pressurised pretreated slurries result in sufficient inhibitor removal and it is the sole step prior to enzymatic hydrolysis, with or without base addition for acid neutralization (Humbird *et al.* (2011) and Kumar *et al.* (2011)). Therefore, the selection of the conditioning strategy depends on the pretreatment, and is thus unique for a process concept. In this case, taking into account that acid impregnated steam explosion results in no major sugar degradation, it is considered that conditioning methods are to be focused on acid neutralization only, and not into removal of possible inhibitory compounds. Thus conditioning refers to the acid neutralization that is necessary in the liquid fraction obtained after pretreatment.

2.4. Lignin recovery

In this design, lignin is intended for energy purposes as mentioned in the previously. Lignin, with low solubility in water at neutral and acidic conditions, can be indirectly recovered by treating biomass with alkali or organic solvents, bringing it into the liquid phase. After separating the solid hollocellulosic fraction, lignin can be precipitated at lower pH. As an alternative approach, lignin can be recovered after enzymatic hydrolysis of the hollocellulose (cellulose and hemicellulose) by a simple solid-liquid separation.

Both lignin recovery alternatives require solid-liquid separations, however lignin recovery prior to hydrolysis implies one extra precipitation step, in addition to recovery of materials in the case of delignification with solvents. Although lignin recovery can be implemented with the alkali pretreatment described in the previous section, high recovery levels require severe conditions or long residence times, which can also imply sugar losses and higher detoxification requirements. On the other hand, delignification after steam explosion has been considered necessary for softwoods like pine. However, studies with SO_2-impregnated bagasse and cornstover have been proved effective, resulting in yields after enzymatic hydrolysis up to 92% of the theoretical glucose yield. This performance has been found common for agricultural residues, whereas softwoods with high lignin content often require delignification (Mabee *et al.*, 2006). Considering that the recovered lignin is intended for supplying energy to the mill, high purities are not necessary. In this context, it is decided that lignin recovery is implemented in the process after enzyme hydrolysis, since it is the simpler alternative and it will be further considered in the design. With this considerations, the process concept for each case is given in the following sections, preceded by the general description of the Base Cases.

2.5. Process alternatives

To select the best process configuration, four cases are considered and compared to a based case that takes into account current practices at palm oil mills. For the development of each case, the following assumptions were considered:

- Mill capacity is 60 tonne FFB/h.
- Plantation area per mill is taken as 18 000 ha based on an annual yields in the range of 19.7 tonne/ha as published by the (Malaysian Palm Oil Board, 2012), taking an annual operation time of 6 000 h/y.
- Availability of EFB, shells, fibres is 23%, 5.5% and 13.5% (w/w) of FFB respectively, as reported by the (Malaysian Department of Environment, 1999).
- For more detailed assumptions and basis of calculations, see Appendix 2.

2.5.1. *Base Case*

The Base Case considers the use of the complete shell and fibre streams for steam and power generation at the mill. The stream of kernels is considered to be exported out of the mill as a product, similar to CPO. The Base Case considers the open-air lagoon system, currently the most common POME treatment method at mills.

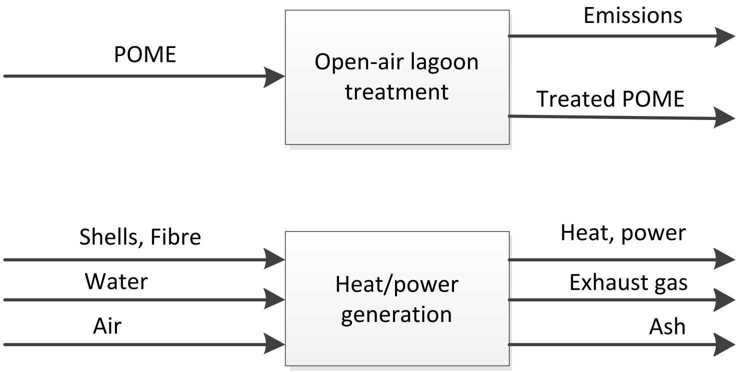


Figure 7. Block diagram of Base Case
POME: Palm oil mill effluent.

2.5.2. *Base Case Plus*

In Base Case Plus biogas is recovered by anaerobic digestion (AD) of the POME and this biogas is used for steam and power generation. As a result, a lower amount of biomass residues is needed for generating the required amount of energy for the palm oil milling.

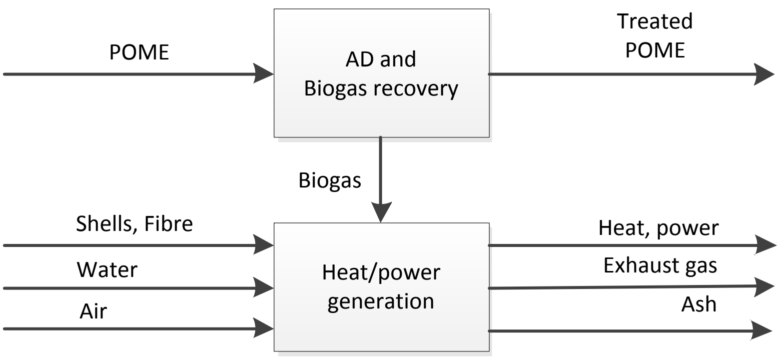


Figure 8. Block diagram of Base Case Plus
POME: Palm oil mill effluent. AD: Anaerobic Digestion

2.5.3. *Case A and Case C*

In Case A and Case C, with biomass derived from palm oil mill operations as main input, an acid catalyst is used for biomass impregnation prior to steam injection into the system. Two streams are obtained from the pretreatment, a liquid fraction containing most of the hemicellulose related sugars, and a solid fraction with the remaining hollocellulose and lignin components. Lime is used for neutralizing the acid left in the biomass, producing small amounts of gypsum as a by-product. Enzymatic hydrolysis is then performed on the separated solids, from which a sugar rich hydrolysate is obtained and mixed with the conditioned liquid fraction derived from the pretreatment. The residual solids (composed of lignin, unreacted hollocellulose and enzymes) are fed into a boiler for heat and power generation, along with biogas obtained from POME anaerobic treatment. The generated steam is a product of the process, but a fraction is also used to cover the pretreatment requirements. The main differences amongst Case A and Case C are the amount of solid biomass entering the plant, where for Case C it is multiplied by the amount of mills taken into account. However, the amount of POME, and thus available biogas, is the same for both situations, because POME derived from a single mill operation is taken into account.

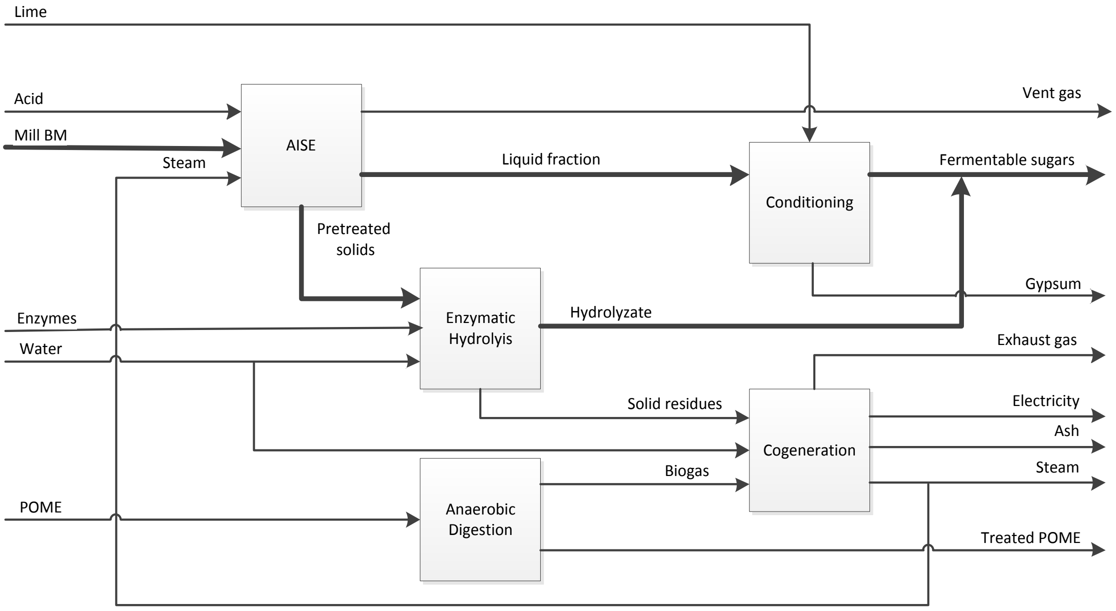


Figure 9. Process concept for Cases A and Case C

AISE: Acid impregnated steam explosion, Mill BM: Empty fruit bunches, shell and fibres, POME: Palm oil mill effluent.

2.5.4. *Case B and Case D*

Similarly to Case A, in Cases B and C, an acid catalyst is used for biomass impregnation prior to steam injection into the system. The process pretreatment and enzymatic hydrolysis are fairly the same as for Case A. However, the introduction of plantation biomass implies an extra step, milling. Plantation biomass, namely OPF and OPT, contain large amounts of mono- and dimeric sugars soluble in a recoverable sap. Thus, prior to pretreatment, these biomass resides are subjected to milling operations in a similar manner as sugarcane, requiring size reduction and milling with imbibition. Specific clarification steps, like sulphitation and liming, are commonly used during sugarcane treatment. However, these are intended for sugar refining to white sugar, which is not considered a requirement in this case. The main differences amongst Case B and Case D are the amount of solid biomass entering the plant, where for Case D it is multiplied by the amount of mills and plantations taken into account. However, similar to Case C, the amount of POME, and thus available biogas, is the same for both situations, where POME derived from a single mill operation is taken into account.

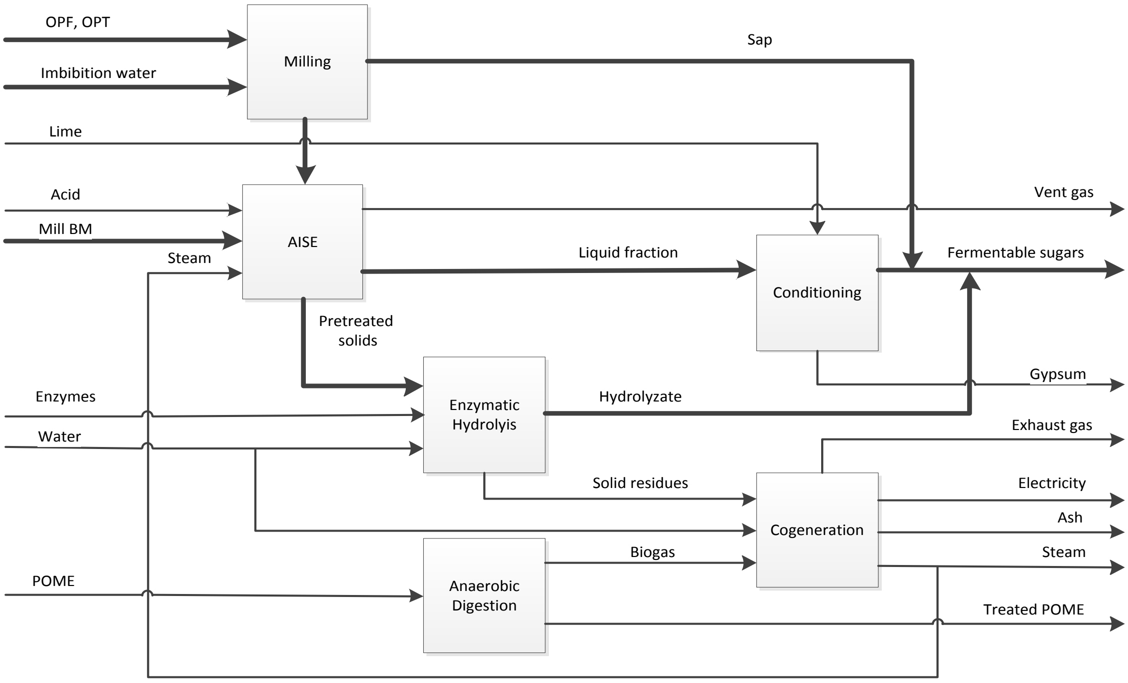


Figure 10. Process concept for Case B and D

AISE: Acid impregnated steam explosion, OPF: Oil palm fronds, OPT: Oil palm trunks, Mill BM: Empty fruit bunches, shell and fibres, POME: Palm oil mill effluent.

2.6. Evaluation

The evaluation of the process alternatives is based on total production costs and their comparison to sugar prices (See Chapter 8 – Economic Evaluation for sugar price considerations). A potential economic margin is calculated taking into consideration the raw material and biomass costs (biomass cost affected by scale is described in detail in Chapter 8.1.1), see Table 4.

Table 4. Raw material cost and potential economic margin, all values are in million RM/y

	BC	BC Plus	Case A	Case B	Case C	Case D
	RM/y					
Raw material cost (excl. biomass)	-	-	3	8	15	41
Biomass cost	5	3	12	49	61	346
Potential income from sugars product	-	-	75	242	374	1210
Potential income from co-products	13	15	10	17	42	71
Potential economic margin	9	11	70	201	339	894

However, the economic potential in Table 4 is just a reference as no overhead cost nor capital charges are taken into account. Thus, in order to estimate production costs, a capital charge representing the annualized contribution of capital investment (CAPEX) was considered. At this point no capital expenditure was calculated as no equipment has been designed yet. At this stage, to have an estimation of CAPEX, installed equipment cost attributed to pretreatment, enzymatic hydrolysis, and milling operations (for sap extraction in Cases B and C) are based on information found in the literature (Eggeman *et al.*, 2005; Carolan *et al.*, 2007; Centro de Gestao e Estudos Estratégicos, 2009; Humbird *et al.*, 2011). For accounting scale effects, equation 1 is used. The scale exponent is taken as 0.7 for pretreatment and enzymatic hydrolysis, and 0.85 for milling operations as reported for these types of operations (Slade *et al.*, 2009; Whitesides, 2012). Raw material cost, excluding biomass, is considered constant at any scale. Biomass cost is considered to change with scale as described in Chapter 8.1.1 – Biomass cost.

$$\mathrm{Cost}_2 := \mathrm{Cost}_1 \cdot \left(\frac{\mathrm{Scale}_2}{\mathrm{Scale}_1} \right)^m \qquad \text{(equation 1)}$$

where:

$Cost_1$ = Cost at the base scale (Scale 1),
$Cost_2$ = Cost at the evaluated scale (Scale 2),
m = Scale index, values can range from 0.3 – 1. For pretreatment and enzymatic hydrolysis it is taken as 0.7, for OPT and OPF milling as 0.85.

The capital expenditure (CAPEX) was estimated based on the Lang factor estimation method. Lang Factoring is a simple method for estimating the capital expenditure of a project when only the main components of the process are known. For this preliminary estimation, the cost of the main process components mentioned above were considered together with a Lang Factor for solid-liquid processing plants, as described by equation 2.

$$\text{CAPEX} := f_{Lang} \cdot \sum C_{eq} \qquad \text{(equation 2)}$$

where:

f_{Lang} = Lang factor, 3.63 for mixed solid-liquid processing,
$\sum C_{eq}$ = Sum of the delivered cost of major items of process equipment.

The change of CAPEX according to scale can be seen in

Figure 11, for processing both mill biomass and mill plus plantation biomass. As expected, when biomass from both mill and plantation is processed, the capital expenditure is larger for the same number of involved mills. Moreover, mill and plantation biomass processing increase more sharply with scale, that is because plantation biomass implies an extra processing step that is less sensitive to economies of scale than the rest, which is sap extraction. However, plantation biomass yields more sugars, making the capital expenditure per annual tonne of sugar produced lower than if only mill biomass is processed. This difference is of course larger at lower scales, as mill biomass processing cost is more sensitive to scales as mentioned before.

Total production costs (TPC) are calculated as a sum of direct production costs (DPC), fixed charges, plant overhead and general expenses. Fixed charges are accounted on a capital charge basis (equation 4 in Chapter 9), while raw materials and produced fermentable sugars change linearly with scale and thus remain constant per amount of produced sugar. Labour and other DPC costs are calculated as function of CAPEX and number of operators with a base wage cost of 90 RM/d (CIDB, 2012), and thus change with scale. Plant overhead expenses, are calculated in relation to labour and maintenance (Seider *et al.*, 2010). General expenses, including administrative and sales charges, remain constant per tonne of produced sugar as they are calculated in relation to sales (Seider *et al.*, 2010). All detailed considerations and numbers are given in Appendix 3 – Process Alternatives.

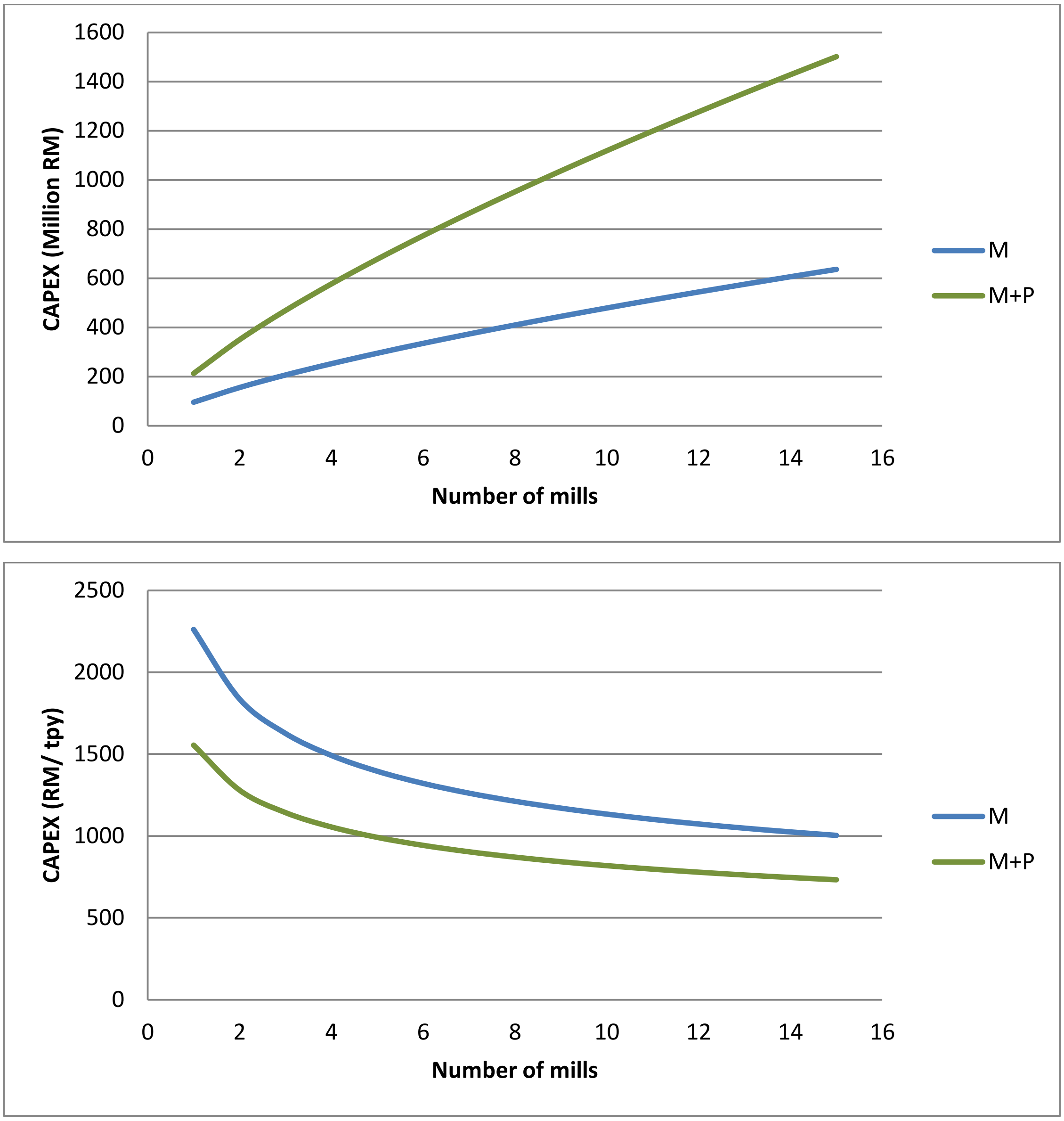


Figure 11. Capital expenditure according to scale

M: Mill biomass for Cases A and C - Empty fruit bunches, shells and fibres. M+P: Mill and Plantation biomass for Cases B and D - Oil palm fronds and oil palm trunks. CAPEX: Capital expenditure, tpy: capacity as tonne of sugar per year.

The change of total production cost with scale can be seen in Figure 12. Economies of scale have a positive impact on production cost only when processing biomass of less than 4 mills for mill-derived biomass and less than 2 mills for mill and plantation biomass. Thus, there is a minimum in production cost, at 4 mill scale for mill biomass (*ca.* 1.2 RM/kg) and at 3 mill scale for mill plus plantation biomass (*ca.* 1.1 RM/kg), points after which the increase in transportation cost has a stronger impact than the benefits of scale related to capital investment. The minimum in production cost is below the long-term contract (LTC) price for raw sugar in Malaysia, 1.77 RM/kg. For comparison purposes, current global prices are in the range of 1.6 RM/kg for raw sugar under the NY contract #12, and 1.4 RM/kg1 for Brazilian sugar (Leahy, 2012; Sugartech, 2012). Although the LTC contract might seem high under current prices, raw sugar prices are sensitive to different phenomena and tend to change in time, for instance the lowest and highest NY raw sugar prices in the period May 2009 – May 2012, were 1.05 RM/kg and 2.32 RM/kg respectively2 (Barchart, 2012), so the LTC price is in a middle point when considering raw sugar prices in the last 3 years.

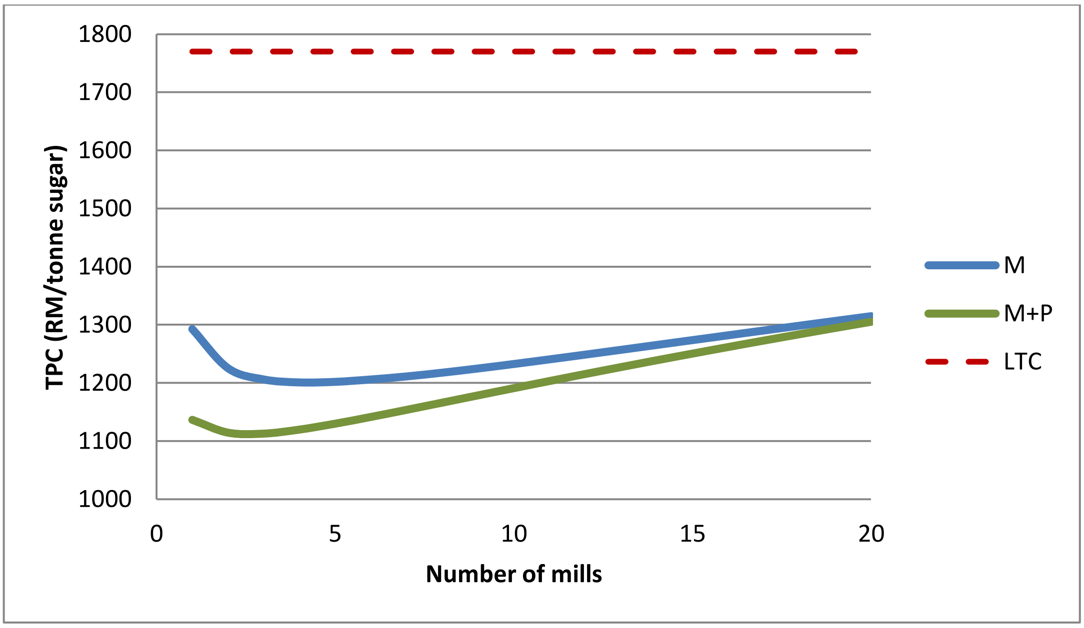


Figure 12. Production costs depending on the amount of biomass derived from n mills.
M: Mill biomass for Cases A and C - Empty fruit bunches, shells and fibres. M+P: Mill and Plantation biomass for Cases B and D - Oil palm fronds and oil palm trunks. CAPEX: Capital expenditure, LTC: Long term contract.

Moreover, the learning or experience effects that come after implementing new processes can be taken into consideration. Learning curves are a way to describe the production cost variations according to cumulative production, due to gained experience and improvements in processes and

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[1] Reported as 19-21 US cent/lb.

2 Prices based on ICE trade values for NY contract no. 11 sugar, with lowest price taken from May 3rd, 2010 and the highest price taken from January 31st, 2011. Current price is taken as 22.73 US cent/lb as of July 16th, 2012.

technology. Brazilian ethanol production cost reduction has been well studied and is described with a global progress ratio (PR, equations 3 and 4) of around 80%, which translated to cost reductions of *ca.* 65% in the 1975 – 2005 period (van den Wall Bake *et al.*, 2009).

$$\mathrm{cost}_{\mathrm{cum}} := \mathrm{cost}_0 \cdot \mathrm{prod}_{\mathrm{cum}}^{\,b}$$

(equation 3)

where:

cost_0 = Reference production cost
$\mathrm{prod}_{\mathrm{cum}}$= Cumulative production
$\mathrm{cost}_{\mathrm{cum}}$= Production cost after certain production time, reaching $\mathrm{prod}_{\mathrm{cum}}$
b = Experience index

$$\mathrm{PR} := 2^{b}$$

(equation 4)

where:

PR = Progress ratio
b = Experience index

Experience or learning curves are based on historical data, however some estimation can be used depending on the process steps. A couple of literature reports discuss cost reductions for transformation of lignocellulosic biomass, in these cases the process is broken down in its different steps and progress ratios are assigned to each section (Slade *et al.* (2009), de Wit *et al.* (2010)). Based on the mentioned literature data, it is expected that both pretreatment and enzymatic hydrolysis will have cost reduction potential in the range of 30%, that is a PR of 0.70, similar to the Brazilian ethanol PR. Considering a cumulative production increase of 15 times (similar to the production increase of Brazilian ethanol in the period 1975 - 2005) the production costs decrease to approx. 35% the original cost of enzymatic hydrolysis and pretreatment capital cost. It is clear that these effects are more visible for mill derived biomass, since sap extraction operations from OPF and OPT are not expected to have a prominent learning effect, Figure 13. Considering these learning effects, the total production costs are decreased below 1 RM/kg when processing biomass derived from less than *ca.* 6 mills.

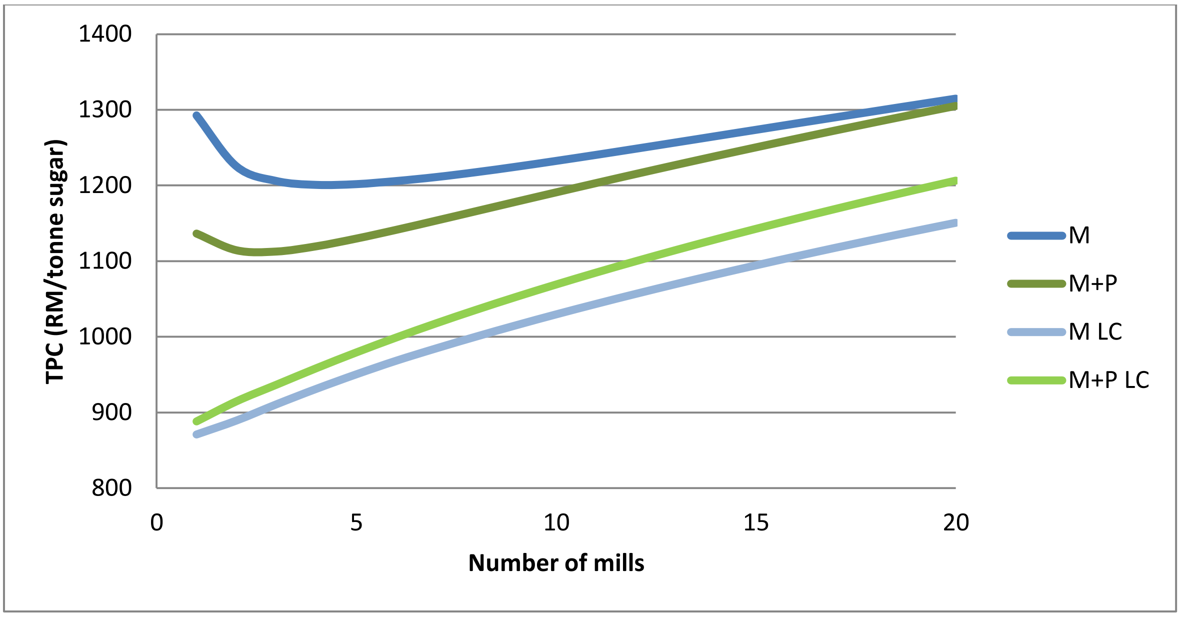

Figure 13. Production costs from variable mills, considering learning effects.
M - Mill biomass - Empty fruit bunches, shells and fibres. M+P: Mill & Plantation biomass - oil palm fronds and oil palm trunks. LC: Considering learning effects.

2.7. Alternative Selection

The primary economic evaluation presented above, is an evaluation of the impacts of economies of scale and biomass transportation costs. This evaluation, including transportation costs, capital charges, overhead cost and general expenses, shows that the production cost of fermentable sugars from palm oil residues greatly depends on scale. The minimum production costs are found at scales of 4 mills for mill biomass in Case C and 3 mills including plantation biomass in Case D. It is found that at higher scales, transportation cost overcome any effect related to economies of scale.

In addition to this, the impact of technological development was estimated considering cost reduction potentials for pretreatment and enzymatic hydrolysis steps. Thus, assuming that 2nd generation fermentable sugars follow the production trend of Brazilian bio-ethanol in its early stages, it can be expected that after 30 years the production costs can go below 1 RM/kg at smaller scales.

Thus, suitable scale has been identified based on economic implications. Based on the results of the evaluation, it is recommended to continue this design project of a 2nd generation palm-based biorefinery that (a) utilises mill and plantation biomass, (b) co-produce fermentable sugars, (c) employs acid catalysed steam explosion treatment, (d) at a scale of 3 mills. For comparison purposes, a parallel design for processing the solid biomass of a 10 mill cluster is going to be presented.

3. Basis of Design

3.1. Description of design

The presented design is part of an exploratory investigation regarding the feasibility of intermediates production from solid residues derived from palm oil processing. Already in the alternative selection section it was shown that smaller scales imply higher economic benefits. Following the recommendations from these preliminary results, the design is focused to the processing biomass available from three oil palm mills and the plantations related to them. However, following discussions with OPBC members, it was agreed that shells conversion is not further considered given its higher lignin content and market price.

Contrasting to the scale findings mentioned above are numerous sugarcane facilities that benefit from economies of scale. It is clear that palm oil processing and supply chain carry different implications to sugarcane operations. Thus, in order to have numbers from scales that are comparable to sugarcane processing, a case regarding the processing of biomass derived from ten oil palm mills is presented as complementary to the three mill case.

3.2. Basic Assumptions

3.2.1. *Plant capacity and seasonal changes*

Calculations are based assuming a mill FFB processing capacity of 60 tonne/h. Plantation area per mill is taken as 18 000 ha based on an annual yields in the range of 19.7 tonne/ha as published by the Malaysian Palm Oil Board (2012), taking an annual operation time of 6 000 h/y. Oil palm fruit bunches are typically harvested all year long. The year-long harvest has a high peak season of four months, typically in the months from July until October. Following similar harvest distribution assumptions as Yeoh (2004), it is considered that 50% of the yearly FFB input can be received on the peak period months, meaning that the mill receives 45 000 tonne/month during peak season and 22 500 tonne/month the rest of the year. These numbers translate to 90 and 45 tonnes/h of FFB for high and low season respectively. However, for the design of the process the average 60 tonne/h per mill is considered, taking that seasonal operating hours can be increased or decreased to cover changes in availability. Thus, availability of EFB and fibres is considered constant at 23% and 13.5% (w/w) of FFB respectively, as reported by the Malaysian Department of Environment (1999).

3.2.2. *Logistics*

3.2.2.1. Biomass transport

Transportation cost of OPF and OPT is based on a plantation area of 18 000 ha per mill and considering an average distance. Assuming a circular plantation area with a mill at the centre, the average distance is taken as 2/3 of the radius, and is thus ca. 5 km. For multiple or *n* mills, the plantation area is considered *n* times the area assigned for feeding a single mill. Additionally, it is considered that EFB and fibres from *n* - 1 mills are to be transported to a central mill. This is considering that biomass derived from operations of for example, 2 out of 3 mills will have to be transported to a 3rd mill. It is taken that distance amongst mills is the average distance from any point in the total plantation area to the centre, see Figure 5. Further considerations on transportation distance and its impact on biomass cost is given in section 8.1 – Raw Material Cost.

3.2.2.2. Plantation biomass storability

Plantations that feed palm oil mills are at different stages of maturity and become ready for replanting at different times. Ideally, replanting is done after around 25 years of productive life when the oil yield has gone lower than 13.6 tonne/ha/y (FELDA, 2012). However, plantations that feed a single mill are not always coordinated to achieve replanting schemes that deliver continuous and constant amounts of trunks. Furthermore, some plantations reach maturities above the recommended 25 years of productive life, especially if palm oil price is high enough to overcome decreased yields. Despite the fact that some larger plantation groups like FELDA are implementing schemes to decrease the number of aged tress in their portfolio, OPT might be available at different times and in different amounts. Thus, it is important that trunks can be stored for buffering and delivering a relatively constant feed into the process.

Felled oil palm trunks are commonly left to rotten on the open-air fields. After felling, trunks can undergo a natural fermentation process that in some countries is enhanced for fermented sap recovery (tapping). Sap tapping is a semi continuous process that gives enough time for bacterial flora to establish themselves in the biomass. This fermentation is thought to be supported on the microbial flora present in the palm and the sap extraction materials that are not commonly cleaned after each tapping (Karamoko *et al.*, 2012). It is thought that lactic acid bacteria first establish themselves and acidify the medium down to pH of 3.5 in the first 3 days of tapping. As the biomass is fermented, the sugar content in the sap extracted from the trunk decreases while organic acids and ethanol increase. (Amoa-Awua *et al.*, 2007).

This type of spoilage of the biomass is not desired as the potential sugars in the biomass are already converted in an uncontrolled manner to ethanol and organic acids, lowering the product potential yield. However, it has been found that when oil palm trunk logs are only left for storage (no tapping activities), the sugar content in the extracted sap is not decreased in the first weeks compared to sap from recently felled trunks (Yamada *et al.*, 2010). On the contrary, in the study by Yamada *et al.* (2010), it was been found that sap extracted from stored logs has increased sugar concentration after 30 – 60 days of log storage. After this time, microbial activity does occur and sugar concentration in the sap is lower. It is thought that starch hydrolytic activity, which is present during the first 30 days of storage, contributes to the increase of sugar concentration in addition to initial cellulose and hemicellulose hydrolysis (Yamada *et al.*, 2010). In view of the previous findings, storage times of 1 - 2 months under shade can be considered if feed buffering is necessary.

3.2.3. *Biomass composition*

Biomass input is composed of separate streams of EFB, Fibres, OPF and OPT. The composition of each of these biomass types is based on a literature study and is presented in section 3.2.6 - Input Streams Definition. Biomass composition is typically reported on cellulose, hemicellulose and lignin basis, sometimes including other components that can be specified as extractives, ash or simply as others. Cellulose, being a well-defined compound, consists of glucose monomers linked by a β (1-4) bond and it is thus considered as a glucan chain. Hemicellulose, on the other hand, can be composed of different polymers containing for instance, xylose, mannose, and arabinose units, with some acetyl side-chains. In the case of EFB composition, hemicellulose has been reported as combined xylan and arabinan (Shamsudin *et al.*, 2012). Since xylan is typically the predominant polymer reported in the hemicellulose fraction of several agricultural residues (around 85% is xylan[3] and the rest is split between mannan, arabinan and galactan), it is considered that xylan is representative of the hemicellulose carbohydrate content. Furthermore, it is known that acetyl groups form part of hemicellulose side-chains, in the range of 4-8 g per 100 g of hemicellulose (Carrasco *et al.*, 2011), thus it is considered that hemicellulose is 94% xylan and 6% acetyl-groups. The extractives component typically encloses other sugars, sugar alcohols and organic acids, thus it is considered as an organic component with the formula CH_2O. In the case of EFB, the extractives fraction was also used to close the mass balance. OPT and OPF composition was calculated back from dry OPF and OPT composition and sap extracted from them, assuming 70% moisture content. Thus, the sugars in the sap are separate from the extractives or undefined fraction reported in the dry biomass analysis.

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[3] Xylan fraction in hemicellulose: 83% in wheat straw, 88% in corn stover (Zimbardi, et al. 2007), 86% in sugarcane bagasse (Carrasco, et al. 2010), 87% in Paja Brava grass (Carrasco, et al. 2011), 84% in switchgrass (Laser, et al. 2009).

3.2.4. *Process model and simulation*

A computer model for the proposed design is built on ASPEN Plus v. 7.3 and is used for simulations regarding the two scale cases. The present model is based on stoichiometric reactions as discussed throughout Chapter 7 – Process and Equipment Design. The native physical properties embedded in the software are used for all components, except for xylan, biomass and enzyme components, which are not found in ASPEN databases. In these cases, new components are defined based on properties described in NREL's 2011 model, see Chapter 4 – Thermodynamic Properties, (Humbird *et al.*, 2011).

3.2.5. *Incoming and outgoing streams summary*

3.2.5.1. Three mill case

Tables 5 and 6 contain a summary of the incoming and outgoing streams to the process, considering biomass derived from three mills processing in total 180 t/h of FFB.

Table 5. Summary of incoming streams, three-mill case

Stream		Quantity	Cost	Total Cost
Name	Number	ktonne/y	RM/tonne	RM/y x 10^3
OPT	< 1 >	540	137	73 751
OPF	< 5 >	660	92	60 440
EFB	< 10 >	248	123	30 483
Fibres	< 12 >	146	70	10 165
Imbibition water	<14 >	508	-	-
Sulphuric acid	< 16 >	10	349	3 559
Water	< 17 >	456	-	-
Lime	< 26 >	7	684	4 925
Water	< 27 >	65	-	-
Enzyme	< 32 >	0	-	23
Water	< 33 >	713	-	-
POME	< 45 >	242	-	-
Boiler Feed Water	< 55 >	780	-	-
Air	< 50 >	60	-	-
Combustion air	< 58 >	1222	-	-
			Raw Material Cost	183 346

Table 6. Summary of outgoing streams, three-mill case

Stream		Quantity	Price	Total Price
Name	Number	ktonne/y	RM/tonne	RM/y x 10^3
Bark	< 3 >	70	137	9 588
Gypsum	< 30 >	12	-2	-19
Fermentable-sugars Product*	< 43 >	357 (3108)	1777	634 570
Emissions	< 52 >	60	-	-
Treated water	< 53 >	521	-	-
Ash	< 59 >	28	-	-
Flue gas	< 60 >	1462	-	-
LP Steam	< 65 >	396	48	19 008
		MWh/y	RM/MWh	
Power**	POWER	72360	283	20 478
			Total	683 625

* In parenthesis is the total mass flow of the sugar product stream. The price of the stream is based on tonne of sugar.

** This is the total produced minus the power required for the equipment, see Appendix 5.

3.2.5.1. Ten mill case

Tables 7 and 8 contain a summary of the incoming and outgoing streams to the process, considering biomass derived from ten mills, processing in total 600 t/h of FFB.

Table 7. Summary of incoming streams, ten-mill case

Stream		Quantity	Cost	Total Cost
Name	Number	ktonne/y	RM/tonne	RM/y x 10^3
OPT	< 1 >	1800	173	312 092
OPF	< 5 >	2202	128	282 702
EFB	< 10 >	828	166	137 652
Fibres	< 12 >	486	113	55 037
Imbibition water	<14 >	1692	-	-
Sulphuric acid	< 16 >	36	349	12 563
Water	< 17 >	1542	-	-
Lime	< 26 >	24	684	16 416
Water	< 27 >	204	-	-
Enzyme	< 32 >	0	-	76
Water	< 33 >	2388	-	-
POME	< 45 >	240	-	-
Boiler Feed Water	< 55 >	2478	-	-
Air	< 50 >	60	-	-
Combustion air	< 58 >	4134	-	-
		18114	Raw Material Cost	816 537

Table 8. Summary of outgoing streams, ten-mill case

Stream		Quantity	Price	Total Price
Name	Number	ktonne/y	RM/tonne	RM/y x 10^3
Bark	< 3 >	234	173	40 572
Gypsum	< 30 >	42	-2	-67
Fermentable-sugar Product*	< 43 >	1194 (10380)	1777	2 121 958
Emissions	< 52 >	60	-	-
Treated water	< 53 >	1200	-	-
Ash	< 59 >	96	-	-
Flue gas	< 60 >	4902	-	-
LP Steam	< 65 >	1194	48	57 312
		MWh/y	RM/MWh	
Power**	POWER	219133	283	62 015
			Total	2 281 790

* In parenthesis is the total mass flow of the sugar product stream. The price of the stream is based on tonne of sugar.

** This is the total produced minus the power required for the equipment, see Appendix 5.

3.2.6. *Input streams definition*

3.2.6.1. Empty fruit bunches (EFB)

During oil extraction, FFB are commonly subjected to a steam treatment to facilitate the subsequent threshing operation in which the fruits are removed from the bunches. The composition of sterilized EFB derived from common mill operation, according to the study of (Shamsudin *et al.*, 2012), is presented in Table 9. It is taken that EFB are available at 23% of FFB, with 68% (w/w) moisture content.

Table 9. composition of Empty Fruit Bunches

Stream Name:	EFB			
Comp.	Units	Specification		Additional Information
		Available (1)	Design (2)	
Moisture	g/kg	-	679	(1) Composition taken from Shamsudin *et al.* (2012). (2) Composition taken from Shamsudin *et al.* (2012), with moisuture content taken from the same source. (3) From mill information given by Syed-Ali (2012)
Cellulose	g/kg	360	116	
Hemicellulose	g/kg	248	80	
Lignin	g/kg	174	56	
Ash	g/kg	78	25	
Others	g/kg	-	45	
Total			1000.0	
Process Conditions				
Temp. (3)	°C		60	
Press.	Bara		1	
Phase	V/L/S		S	

3.2.6.2. Oil palm fronds (OPF)

The available amount of OPF is based on frond availability at pruning and replanting, 10.4 t/ha/y and 14.4 t/ ha respectively (Yusoff, 2006), and is 0.55 tonne per tonne of FFB. However, only the petiole part (1/3) of the frond is taken into account as the leaflet is considered for nutrient recycle at the field (Agensi Inovasi Malaysia, 2011). Studies on the composition of the OPF, with 70% moisture, have been focused on either the dried fibres or the sap obtained from it (Zahari *et al.*, 2012). The composition of the fronds fibres and the sap is presented in Table 10, with data from (Zahari *et al.*, 2012).

Table 10. Composition of Oil Palm Fronds (OPF)

Stream Name:		**OPF**		
Comp.	Units	Specification		Additional Information
		Available (1)	Design (2)	
Moisture	g/kg	-	700	(1) Oil palm frond fibre composition reported by Zahari *et al.*, (2012). (2) Composition based on (1) assuming moisture content and sap composition as reported by Zahari *et al.* (2012). (3) Taken as ambient temperature.
Cellulose	g/kg	417	125	
Hemicellulose	g/kg	164	49	
Lignin	g/kg	155	47	
Glucose	g/kg	-	41	
Fructose	g/kg	-	2	
Sucrose	g/kg	-	15.9	
Others	g/kg	264	20.8	
Total			1000.0	
Process Conditions				
Temp. (3)	°C		25	
Press.	Bara		1	
Phase	V/L/S		S	

3.2.6.1. Fibres

Fibres available at the mill account for 13.5% (w/w) of FFB. The composition of this material varies among several literature sources, and thus it is taken as the average of different reported values presented in Table 11 (Kumar *et al.*, 2010).

Table 11. Composition of fibres

Stream Name:		**Fibres**		
Comp.	Units	Specification		Additional Information
		Available (1)	Design (2)	
Moisture	g/kg	-	400	(1) Dry weight basis from literature review as presented by Kumar *et al.* (2010). (2) Composition taken from (1), assuming typical moisture content reported by Schmidt, (2007). (3) Taken as ambient temperature.
Cellulose	g/kg	364	218	
Hemicellulose	g/kg	235	141	
Lignin	g/kg	281	169	
Ash	g/kg	46	28	
Extractives		74	44	
Total			1000.0	
Process Conditions				
Temp. (3)	°C		30	
Press.	Bara		1	
Phase	V/L/S		S	

3.2.6.2. *Oil palm trunks (OPT)*

The availability of OPT is 75.5 DW tonne per replanted hectare (Yusoff, 2006). Based on total plantation area of 18 000 ha and replanting after 25 years, this yields 0.15 tonnes of dry trunk per tonne of FFB. Similarly to OPF, studies on the composition of the OPT with typical moisture contents of 70%, have been focused on either dried fibres or sap obtained from it. The composition of OPT based on the composition of both sub-products is presented in Table 12, with on data from (Paterson *et al.*, 2009) and (Kosugi *et al.*, 2010).

Table 12. Composition of oil palm trunks (OPT)

Stream Name:		**OPT**		
Comp.	Units	Specification		Additional Information
		Available (1)	Design (2)	
Moisture	g/kg	-	700	(1) Oil palm trunk fibre composition reported by Paterson *et al.* (2009). (2) Composition based on (1) assuming moisture content and sap composition as reported by Kosugi *et al.* (2010). (3) Taken as ambient temperature.
Cellulose	g/kg	412	110	
Hemicellulose	g/kg	344	92	
Lignin	g/kg	171	46	
Glucose	g/kg		39	
Xylose	g/kg	-	1	
Fructose	g/kg	-	2	
Sucrose	g/kg	-	2	
Ash	g/kg	34	9	

Total		1000.0	
Process Conditions			
Temp. (3)	°C	25	
Press.	Bara	1	
Phase	V/L/S	S	

3.2.6.3. Palm oil mill effluent (POME)

POME is the waste water effluent derived from common mill operations. It is mainly composed of steriliser condensate, oil clarification waste water and hydrocyclone waste water. It is considered that POME is produced at a rate of 0.67 t/t of FFB (Schmidt, 2007). The composition of this liquid stream is relatively similar for different mills, and is presented in Table 13 with data from the Malaysian Department of Environment, 1999. In the design it is assumed that residual solids have been screened off.

Table 13. composition of Palm Oil Mill Effluent (POME)

Stream Name:		**POME**		
Comp.	Units	Specification		Additional Information
		Available (1)	Design (2)	
Oil and grease	g/kg	0.150 - 18	6	(1) Ranges given by the Malaysian Department of Environment, (1999). (2) Mean averages given by the Malaysian Department of Environment (1999). (3) Typical POME discharge temperature. (4) BOD: Biochemical Oxygen Demand, COD: Chemical Oxygen Demand. (5) The mean COD 50 g/L reported by the Malaysian department of Environment is taken as basis for biogas emissions.
BOD (4)	g/kg	10 - 44		
COD (5)	g/kg	16 - 100	50	
Total Solids	g/kg	11.5 - 79		
Volatile Solids	g/kg	9 - 72		
Water			954	
Total			1000.0	
Process Conditions				
Temp. (3)	oC		80	
Press.	Bara		1	
Phase	V/L/S		L	

4. Thermodynamic Properties

Properties used for calculations are based on native properties in ASPEN Plus databases. However, for enzyme, biomass and xylan components the properties were taken from NREL 2011 report (Humbird, *et al.* 2011). See Table 14 for detailed information on properties for non-native ASPEN Plus components.

Table 14. Input Properties for individual components

Compound	ASPEN name	Property	Quantity	Units	Reference
Enzyme	ENZYME	MW	22.8398	-	Humbird *et al.*, 2011
		DHSFRM	-73 713 712	J/kmol	
		VSPOLY	(1) 0.015 (6) 298.15	m^3/kmol	
		CPSP01	(1) 35533 (7) 298.15	J/kmol K	
Fructose	FRUCTOSE	-	-	-	As glucose, native ASPEN Component
Xylan	XYLAN	MW	132.1170	-	Humbird *et al.*, 2011
		DHSFRM	-761 906 107	J/kmol	
		VSPOLY	(1) 0.086 (6) 298.15	m^3/kmol	
		CPSP01	(1) -9529.9 (2) 547.25 (7) 298.15	J/Kmol K	
Lignin	LIGNIN	-	-	-	As vanillin, native ASPEN Component
Biomass	BIOMASS	MW	23.2380	-	Humbird *et al.*, 2011
		DHSFRM	-97 068 842	J/kmol	
		VSPOLY	(1) 0.016 (6) 298.15	m^3/kmol	
		CPSP01	(1) 35910 (7) 298.15	J/kmol K	
Extractives	EXTRACT	-	-	-	Organic extractives, as glucose native ASPEN Component

The thermodynamic model selected for the simulation is the non-random two-liquid model (NRTL). This method is chosen as it is recommended for polar, non-electrolyte systems at moderate pressures, and is typically used in chemical process applications including solid biomass conversions. This property method is used globally in the simulations as it was considered accurate enough for the present study.

5. Process Structure and Description

5.1. Process Flow Scheme

Process flow diagrams are given in Appendix 4. A simplified diagram of the process, with an overview of the different process areas is shown on Figure 14.

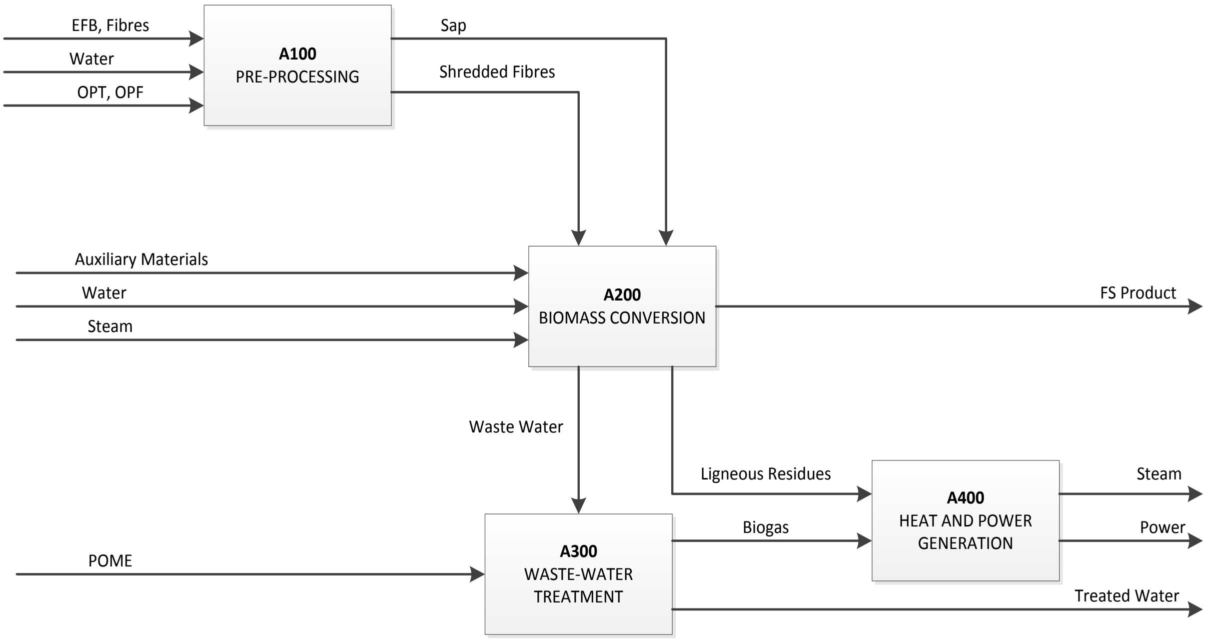

Figure 14. Simplified process scheme

FS Product – Fermentable Sugars Product, POME: Palm Oil Mill Effluent, OPT: Palm Oil Trunks, OPF: Palm Oil Fronds, EFB: Empty Fruit Bunches.

5.2. Area 100 – Pre-processing

5.2.1. *Biomass preparation*

The required pre-processing for the biomass depends on the type of biomass and its intended use. Pre-processing of OPT and OPF is required for recovering the sugar-rich palm sap, while EFB are pre-processed to deliver adequate and homogenous biomass into the pretreatment reactor.

The basal part of the oil palm frond can be considered similar to sugar cane, in both dimensions and composition. Typically, the first step in sugarcane preparation is cutting the cane with knife blades arranged in 2 or 3 sets to provide a uniformly chopped material for shredding (Hugot, 1986). However, FELDA has run preliminary experiments where cutting operations were found to be unnecessary due to the less harsh frond structure (Syed-Ali, 2012). Thus, fronds can be directly passed through shredders to prepare for sap recovery. The function of this shredding equipment is to break the cells and disintegrate the structure. As a result, the material leaves as long thread fibres. After shredding, biomass is ready for feeding into the milling tandem.

As tree trunks, the OPT stream needs to first undergo through debarking step, after which it can be shredded and fed into a mill. Oil palm trunks can be pre-processed on site with a debarker unit, delivering bark for plywood producers and the oil palm trunk core for further processing. However, the stability of trunks with storage is expected to be lower if they are debarked, considering that the sugar rich core would be exposed to the environment. Thus, debarking should be done at the facilities prior to the rest of the pre-processing. It is then considered that OPT can be bundled and transported to the biorefinery for buffering storage, and are available for debarking and shredding prior to milling.

It is considered that EFB can be processed in the same type of equipment as OPF for shredding. Two-separate lines are considered beneficial due to logistics, one processing EFB and the other one OPF to be delivered for further processing. Fruit fibres available from oil extraction, with thickness of a couple of millimetres, are not considered in primary pre-processing.

5.2.2. *OPF and OPT milling*

Sap recovery from oil palm and trunks has been reported in the literature, yielding 0.5 and 0.65 tonne/tonne of OPF and OPT respectively using typical laboratory pressing equipment product (Kosugi *et al.*, 2010; Zahari *et al.*, 2012). In perspective, during sugarcane milling it is standard practice to include imbibition as part of the processing in order to recover most of the sugars present in the biomass. As the milled biomass holds moisture in the range of 50%, imbibition water is used to replace the juice retained in the biomass, thus reaching sugar recoveries of about 95% w/w (Hugot, 1986). Water utilization is typically mentioned in the range of 20 – 35% on cane wet weight basis; sugar mills in Brazil, Mauritius and India report water use as 167 – 223% on fibres basis (Johnson *et al.*, 2012). It is considered that OPT and OPF can undergo a similar milling process with imbibition to reach a high sugar recovery. The sap extracted from the milling operation is screened to removed solids and particles.

5.3. Area 200 – Biomass Conversion

5.3.1. *Pretreatment*

As mentioned in previous sections, prior to the enzymatic hydrolysis of cellulose, a pretreatment step is necessary. In section 2 – Process Definition, an analysis of different pretreatment technologies was carried out. After comparing different technologies it was decided that acid catalysed steam explosion was the best alternative.

During this processing step the structure of the lignocellulosic material is opened due to the pressure changes, at the same time most hemicellulose is solubilized due to acid groups. As little or no information was found regarding steam explosion of OPF, OPT or EFB, the selection of conditions for this step is based on other biomass materials, particularly agriculture residues such as corn stover, switchgrass and sugarcane bagasse.

As the biomass is delivered to the pretreatment section by carriers, it enters a preheater. The equipment is fed with low pressure (LP) steam available from the combustion of solid waste streams in the CPH system. The wet biomass is transported through screw conveyors into the high pressure pretreatment reactor. A 2% H_2SO_4 solution is injected to reach loadings of 22 kg/tonne of dry biomass. Medium pressure (MP) steam is injected into the reactor, which is maintained at 190 °C. The hot slurry is flashed to 1 bar; the resulting flash gases are composed of volatile compounds like furfural and acetic acid.

The slurry from the pretreatment reactor is acidic and must be conditioned prior to enzymatic hydrolysis. In section 2 it was discussed that the conditioning method is based on acid neutralization with lime. In order to better condition the slurry, it can be separated into liquid and solid fraction to separate the formed gypsum slurry and avoid interference with the enzymatic reaction. A screw press equipment is used to separate the liquid fraction from the solids.

The low pH liquid is neutralized with lime, to reach a pH of 5. The formed gypsum is removed with a rotary filter and exported as waste. The conditioned filtrate is used to dilute the cellulosic solids prior to entering the enzymatic hydrolysis.

5.3.2. *Enzymatic hydrolysis*

In order to minimize the dilution of the resulting fermentable sugars and minimize the water use, it is decided to run the enzymatic hydrolysis at total solid loadings of 20% w/w (soluble plus insoluble). Although the conditioned filtrate is mixed back with the cellulosic solids, some dilution water is still required. It is considered that the enzyme solution is prepared with the required dilution water. The enzymatic hydrolysis takes place with enzyme loading of 4% w/w at 50 °C.

It is considered that the hydrolysis can run in parallel reactors. The reactors are vertical towers in which the biomass flows downwards as it is continuously hydrolysed, similar to those described in 2011 NREL report (Humbird *et al.*, 2011). After the hydrolysis of cellulose, the stream is separated into a solid and a liquid fraction. The solids are exported to the heat and power generation area, while the liquid containing lignocellulosic-derived sugars is combined with the sap from milling operations. The sugar rich stream is then subjected to high temperatures for short retention time to limit the growth of microbial pollutants and prevent spoilage.

5.4. Area 300 – Waste water treatment

Waste water is originally composed of POME from the adjacent mill. Combined with POME is the condensate from the pretreatment section. The combined waste water streams enter an anaerobic digestion treatment; biogas is recovered from the digester as the COD content is removed from the water stream. The outflowing water stream, containing remaining organics, is stabilized in an aerobic step. This section is based on previous results obtained from POME treatment with biogas recovery.

5.5. Area 400 – Heat and Power Generation

The solid stream separated from the enzymatic hydrolysate is combined with the overflow sludge from the waste water area. The wet solid materials are preheated with heat recovered from the flue gases. The combustion heat is then used to raise steam. Fresh water is preheated and combined with recovered hot condensed steam used in the process. Superheated steam is raised and passed through an extraction turbine. Two different steam qualities are recovered from the turbine: (1) Medium pressure (MP) steam to feed the pretreatment system, and (2) low pressure (LP) steam to feed both the proposed process and the adjacent palm oil mill.

5.6. Process Stream Summary

The process stream tables can be found in Appendix 9.

6. Mass and Heat Balances

Mass and energy balances were solved using ASPEN Plus 7.3, MATLAB and Microsoft Excel. The considerations for the mass balances and biomass composition and conversion are given in Section 3 and Section 8. Energy flows are presented based on enthalpy estimations solved using ASPEN Plus. Enthalpy estimations are based on the sum of 1) enthalpy change involved in reacting the elements at 25 °C and 1 atm in their reference state conditions to form the compound at the same temperature and pressure (enthalpy of formation), 2) enthalpy change involved in taking the compound from the reference temperature to the system temperature, and 3) enthalpy chance involved in taking the compound to the system pressure and state (enthalpy departure).

The complete mass balances for each equipment and the process are given in Appendix 5. Overall mass balances close to 100%, whereas energy balances carry in 0.65% error. The small difference between incoming and outgoing energy flows is mainly introduced by rounding-off temperatures and enthalpies from ASPEN Plus simulations to Microsoft Excel sheets and by errors introduced with numerical solving of process design specifications in ASPEN Plus.

7.Process and Equipment Design

7.1. Three-mill case

7.1.1. *A100 – Pre-processing*

*7.1.1.1. OPF preparation **<A101>***

Considering that sugarcane preparation for milling is well established, it can be considered as a model for preparation of OPF. Typical sugarcane preparation operations include the use of knife-cutters and shredders. Cane can be cut with knife blades arranged in 2 or 3 sets to provide a uniformly chopped material. Typical speed of rotation for the knife blades are in the range of 400 to 720 rpm, with 500 rpm as the most common value (Hugot, 1986). However FELDA Bhd research division has tested fronds with sugarcane processing equipment, finding that due to the fronds softer outer layer (as compared to sugarcane), a cutting step is not necessary (Syed-Ali, 2012).

Thus, delivered fronds are passed directly through shredders to prepare for sap recovery. The function of this equipment is to break the cells and disintegrate the structure. It is reported that total installed power for sugarcane preparation is in the range of 20 kW/tfh (tonne of fibre per hour) for the knives and 15 – 20 kW/t.f.h. for the shredder (Hugot, 1986). Considering the average values and an input of 660 ktonne/y of OPF with 70% moisture content (assuming tonne of dry weight as tonne of fibre), this results power requirement of 578 kW for OPF shredding, Table 15.

Table 15. Specifications for OPF shredder

Specification			Unit	A101
Input:	Cut OPF	< 5 >	tonne/h	110
Output:	Shredded OPF	< 6 >	tonne/h	110
Power consumption			kW/dry tonne	17.5
Power			kW	578

*7.1.1.2. OPT preparation **<A102> <A103>***

Debarking of felled trees is a well-studied operation, for which large scale equipment is available. In order to estimate the mass flows around the debarking operation, it is necessary to know the amount of bark that can be removed from the trunk. The bark weight contribution to the tree trunk (above-ground part of the tree without crown and branches) varies for different types of trees. No data was found for oil palm trees, but it was found that bark contributes with 17.20%, 13.27%

and 13.14% for maple, pine tree and Douglas-fir trees (Briggs, 1994; Shmulsky *et al.*, 2011). Taking this bark fraction values into account, it is assumed that bark in the palm trunk contributes to 13% w/w of the estimated available trunk biomass. Thus, the available output for pre-processing is 78.30 tonne/h of debarked trunk. Although these types of operations are not the most common practice, they already take place for production of plywood from OPT, see Figure 15.

Figure 15. OPT transportation and debarking operations, taken from Yutaka et al. (2007)

Debarking processing is common in the wood industry, with three main technologies: ring style debarkers, cradle debarkers and enzyme assisted debarkers. Power consumption is taken as for the more common dry ring debarkers, with energy demand of 0.025 GJ/tonne debarked log (Clark *et al.*, 2007), which would result in 544 kW.

Several debarker models with different capacities are available in the market. For large capacities, double rollers peeling equipment is necessary. Complete OPT trunks as delivered to the mill can be fed into the rotary drum, in which the material is peeled one log at the time. Considering that in average there are 90 tonne/h of OPT, three 12 meter roller derbarker units can be used, see Appendix 7 – 1.1.1 for more detail and Table 16 for considered specifications. As the available amount of OPT is not constant, it is considered that 3 units is a good option for buffering variable inputs.

Shredding equipment for OPT has been developed specifically for sap recovery applications. The equipment takes in the debarked trunk core and shreds it directly into chips ready for milling, with no need for separate knives. In a first stage the equipment rolls the palm trunk core and delivers it to a cutting section where it is received by a rotary cutter (Murata *et al.*, 2009). No energy consumption was found for this of equipment, thus typical values for shredding equipment as given by Hugot (1986) are considered. Since the OPT material to be shredded is in the form of whole trunk core, the highest value in the range of the typical values is taken, i.e. 20 kW/tfh. The resulting power requirement is 470 kW for the 78.3 tonne/h of debarked trunk, Table 17.

Table 16. Specifications for OPT debarking equipment

Specification			Unit	A102
Input	OPT	< 1 >	tonne/h	90
Output	Bark Debarked trunk	< 3 > < 2 >	tonne/h tonne/h	12 78
Available capacity			tonne/h	30
Units needed			tonne/h	3
Bark removal from total trunk			w/w	0.13
Power			kW	544

Table 17. Specifications for OPT shredder

Specification			Unit	A103
Input	Debarked trunk	< 2 >	tonne/h	78
Output	Shredded trunk	< 4 >	tonne/h	78
Power consumption			kW/dry tonne	20
Power			kW	470

7.1.1.3. OPT and OPF milling ***<A104>***

Milling of OPT chips obtained as described in the previous section has been performed using mills with 2 sets of three rotary-hydraulic-press rollers. The rollers used for OPT chips have been design for high recoveries (large grooves and low-cut chevron), and the highest sap recovery achieved is in the range of 80% (Murata *et al.*, 2009). This could probably be improved by using more sets of rollers. Typical mill configurations used for sugar cane processing are arranges of 3 to 7 three-roller sets, yielding around 95% of sugar recovery with aid of imbibition. Thus, for the design of the milling tandem, typical characteristics used for sugarcane processing are considered, which can also be taken for milling shredded OPF.

Based on equations and assumptions given in Appendix 7 – 1.1.2, the milling tandems consist of 14 rollers, with one crusher and 4 mill sets, and maximum processing capacity of 211 tonne/h. The total power requirement for the milling tandems is 4 144 kW. Summarised specifications are given in Table 18. Imbibition is considered necessary to reach high sugar yields, it is estimated that by using imbibition water as 1.5 tonne/tonne dry biomass, 93% of the sugars can be recovered in the sap.

Table 18. Milling Tandems Specifications

Specification			Unit	A104
Input	PBM	< 7 >	tonne/h	188
	Water	< 14 >	tonne/h	85
Output	Sap	< 9 >	tonne/h	178
	PBMF	< 8 >	tonne/h	95
Fibre fraction			w/w	0.3
Roller speed			rpm	6
Number of rollers (incl. crusher)			-	14
Tandems			-	4
Power consumption			kW/tph	22
Total Power			kW	4 144

7.1.1.4. EFB shredding ***<A105>***

Although EFB have already undergone a type of pre-processing (sterilizing and threshing), it is necessary to reduce the size of the biomass. This is to 1) make the biomass material more susceptible for conversion, 2) avoid physical clogging of the pretreatment reactor, 3) deliver homogenous input to the pretreatment reactor with all biomass shredded into fibres. Similarly to OPT preparation, there is some equipment available specifically for EFB shredding. MSHK Engineering Sdn. Bhd. trades an EFB shredder with capacities as high as 7 tonne/h, which would result in the need of 6 shredding units to process the almost 42 tonne/h of EFB available from 3 CPO mills. Considering the energy consumption for shredding operations as mentioned by Hugot (1986), the power requirement is 233 kW.

Table 19. EFB Shredder specifications

Specification			Unit	A105
Input	EFB	< 10 >	tonne/h	41
Output	Shredded EFB	< 11 >	tonne/h	41
Power consumption			kW/dry tonne	18
Power			kW	233

7.1.2. *A200 – Biomass Conversion*

7.1.2.1. Pretreatment <E201> <R201>

Acid catalysed steam explosion has been tested with few acid catalyst. Research has been mostly focused on H_2SO_4 and SO_2, and these two options are considered. After a review on these two catalysts, see Appendix 7 – 1.1.3, sulphuric acid is selected for pretreatment at 190 °C. The equipment set-up for pretreatment is based on NREL's 2011 proposed set-up, however as the available biomass has high moisture content, a receiver with water addition is not considered necessary. Thus, the pretreatment equipment considered for this process is composed of: 1) a preheater in which steam is injected to raise the wet biomass temperature to 100 °C, increasing at the same time the moisture content of the biomass entering the pretreatment reactor, and 2) a horizontal high pressure reactor at 190 °C into which the biomass and high pressure steam are added.

Pretreatment under acidic conditions results in high hemicellulose solubilisation, while most cellulose remains in the solid fraction. Based on experimental results given in Appendix 7, it is considered that xylose obtained from xylan after pretreatment is 90% of the theoretical yield. Furfural formation on H_2SO_4 impregnated steam explosion treatments is not reported abundantly in the literature. Results under similar conditions with H_2SO_4 show formation of furfural as 2.8% w/w of the total xylan, while steam explosion with SO_2 have resulted in 0.52% w/w furfural release (Zimbardi *et al.*, 2007; Carrasco *et al.*, 2011). In comparison, furfural formation has been reported in the range of 5% of total xylan with dilute acid treatments with sulphuric acid. The lower degradation of xylose during steam explosion pretreatment is reasonable considering that the process is carried during shorter retention times. It is thus considered that 2.5% of xylan is degraded to furfural.

Cellulose hydrolysis is minor during acid pretreatments. Reported values with dilute acid treatment are typically in the range of 5%, with similar values reported by Carrasco *et al.* (2011) for SO_2 catalysed steam explosion. It is considered in this design that 5% of the glucan is converted to glucose. Furthermore, glucose degradation to hydroxy-methyl-furfural (HMF) has been reported as low as 0.17% of the total glucan or none for similar acid catalysed steam explosion conditions (Zimbardi *et al.*, 2007; Carrasco *et al.*, 2011). In comparison, NREL dilute acid pretreatment model, with 5 min. at 190 °C plus 30 min. at 130 °C takes 0.3% w/w of glucan to HMF (Humbird *et al.*, 2011). Based on the previous observations, HMF formation is considered negligible and no glucose degradation reaction is entered in the pretreatment model. Furthermore, based on the study by Bower *et al.* (2008) and considering the short retention time, it is assumed that all residual sucrose in OPT and OPF is hydrolysed to glucose and fructose, without further degradation.

Deacetylation of hemicellulose side-chains is one of the common outcomes of acid pretreatments. As result, dilute acid pretreatment models typically consider 100% conversion of the acetyl side chains in the biomass to acetate. Acetyl groups content in the biomass is related to the hemicellulose side chains, with common acetyl-/hemicellulose ratios in agricultural residues in the range of 4-8% (Carrasco *et al.*, 2011). Deacetylation during acid catalyzed steam explosion has been reported in different ranges. Zimbardi *et al.* (2007) reported up to 3g of acetic acid obtained from 100g of dry corn stover with H_2SO_4, while 0.22g for grass and 2.3 g for bagasse have been reported using SO_2 at 190 °C (Carrasco *et al.*, 2010; Carrasco *et al.*, 2011). It is considered that 100% of the acetyl content estimated in the biomass is released as acetic acid, i.e. 1.4 g/g of dry biomass. As for lignin degradation products, typical values with dilute acid pretreatments are in the range of 1 - 5% solubilisation of original lignin in the biomass (Laser *et al.*, 2009; Humbird *et al.*, 2011; Kumar *et al.*, 2011), whereas the only value found in the literature for AISE is below 1% (Carrasco *et al.*, 2011). It is thus considered that 1% of solid lignin is solubilised during pretreatment.

Table 20. Reactions considered in pretreatment simulation

Reaction	Conversion
$(C_6H_{10}O_5)_n + n\ H_2O \longrightarrow n$ Glucose	5%
$(C_5H_8O_4)_n + n\ H_2O \longrightarrow n$ Xylose	90%
Xylose $\longrightarrow$ Furfural + 3 H_2O	3%
Acetate $\longrightarrow$ Acetic Acid	100%
Lignin $\longrightarrow$ $\text{Lignin}_{\text{soluble}}$	1%
Sucrose $\longrightarrow$ Glucose + Fructose	100%

Table 21. Pretreatment system specifications

Specification	Units	E201	R201
Incoming streams	-	< 13 > < 68 >	< 15 > < 18 > < 62 >
Outflow streams	-	< 15 >	< 20 >
Flow	tonne/h	172	302
Residence time	min	10	2
Design volume	m3	24	14
Steam requirement	tonne/h	11	52
Temperature	°C	100	190

7.1.2.2. Fibres separation ***<S202> <S204>***

The slurry from the pretreatment reactor is acidic and must be conditioned prior to enzymatic hydrolysis. In section 2 it was discussed that the conditioning method is based on acid neutralization with lime. The pretreated slurry can be separated into liquid and solid fraction to separate the solid gypsum and avoid interference with the enzymatic reaction. Equipment for dewatering of biomaterial is commonly used in waste water and pulp and paper industries, and a review of several alternatives is presented in Appendix 7 – 1.4. A screw press that delivers a biomass cake with 65% w/w moisture is considered in the design for separating the solid fraction from the pretreated slurry < 21 >. Furthermore, a separate screw press is considered for the ligneous slurry remaining after the enzymatic hydrolysis in reactor R203. For the design specifications, similar conditions are considered as for processing the pretreated slurry, see Table 22.

Table 22. Screw Press specifications

Specification	Units	S202	S204
Incoming steams	-	< 21 >	< 36 >
Outflow streams	-	< 23 > < 24 >	< 37 > < 38 >
Flow	tonne/h	253	380
Max. capacity	tonne/h	140	140
Units		2	3
Av capacity used	tonne/h	127	127
Power per unit	kW/unit	97	97
Total power	kW	388	583

7.1.2.3. Conditioning ***<R202> <S203>***

In Section 2 several conditioning alternatives are overviewed and lime is selected for acids neutralization. It is considered that $Ca(OH)_2$ is used for increasing the pH to 5 that is the optimal pH for enzymatic hydrolysis enzymes, as mentioned in Section 2. In this step the equilibrium reactions in Table 23 are considered to determine the amount of lime that is required to reach the target pH. The equilibrium reactions, in combination with mass and charge conservation equations, were solved using MATLAB symbolic solve function for non-linear equations. As the resulting $Ca(CH_3COO^-)_2$ concentration is minimal, it is not considered in the ASPEN model. Phenolic compounds and furans are typically considered to precipitate when conditioning is based on overliming. However, as the pH in this case is only increased to 5 (not up to 10 and then down to 5 as overliming would imply), no furan or phenolic compounds precipitation is considered. The decision to just increase the pH to hydrolysis conditions is based on the relatively low concentration of degradation products, which is below levels considered detrimental for microbial activity, thus avoiding larger amounts of gypsum slurry.

Table 23. Reactions considered in conditioning step

Reaction	k
$H_2SO_4 \longrightarrow H^+ + HSO_4^-$	1×10^3
$HSO_4^- \longrightarrow H^+ + SO_4^{2-}$	1×10^{-2}
$CH_3COOH \longrightarrow H^+ + CH_3COO^-$	1.3×10^{-5}
$Ca(OH)_2 \longrightarrow Ca^{2+} + 2OH^-$	5.02×10^{-6}
$CaSO_4 \longrightarrow Ca^{2+} + SO_4^{2-}$	4.93×10^{-5}
$Ca(CH_3COO^-)_2 \longrightarrow Ca^{2+} + CH_3COO^-$	1×10^2

Residence times for neutralizing operations vary from 10 – 20 min for waste waters. To prevent downstream pH changes associated with post-neutralization reactions, it is decided to take 20 minutes as residence time for the design (Goel *et al.*, 2005), see Table 24 for considered specifications.

Table 24. Conditioning reactor specifications

Specification	Units	R202
Inflowing streams	-	< 25 > < 28 >
Outflow streams	-	< 29 >
Flow	tonne/h	159
Residence time	min	20
Design volume	m3	65
Operational volume	m3	51
Tip speed for agitator	m/s	3.3
Power	kW	10

A rotary drum filter is considered for removal of the solids formed during conditioning, in the form of $CaSO_4$ slurry. For the simulation, it is assumed that gypsum particle size is in average 40 μm, and that the cake compressibility factor is 0.5 based on common values ranging from 0.2 – 0.8 (McCabe *et al.*, 2001). It is considered that the rate of revolution is 2 rpm to reach solids content in the cake to 75% of the cake mass. The equipment design is based on a design filtration rate as 323.3 $kg/h.m^2$, estimated for drums with maximum effective submergence of 30%, delivering a cake with 25% moisture and thickness of 7.5 mm, see Table 25 for specifications. The total area required is 492 m^2 thus 3 units of 180 m^2 required considering reported maximum drum filter areas of 186 m^2 (Perry *et al.*, 1997).

Table 25. Rotary drum filter specifications

Specification	Unit	S203
Inflowing streams	-	< 29 >
Outflow streams	-	< 30 > < 31 >
Flow	tonne/h	159
Units required	-	3
Total power	kW	393

*7.1.2.4. Enzymatic hydrolysis **<R203>***

As mentioned in Section 2, it is decided to acquire commercial enzyme preparations. Typical commercial preparations for non-starchy biomass contain cellulases and in some cases hemicellulases, resulting in soluble gluco- and xylose- oligomers and monomers. Based on commercial enzymes currently available in the market, the achievable conversions are presented in Table 26.

Table 26. Reactions considered in the enzymatic hydrolysis step

Reaction	Conversion*
$(C_6H_{10}O_5)_n + n\ H_2O \longrightarrow n\ C_6H_{12}O_6$	90%
$(C_5H_8O_4)_n + n\ H_2O \longrightarrow n\ C_5H_{10}O_5$	90%

*Conversion based on remaining glucan and xylan in the pretreated biomass.

In order to favour the economics of the process, working at high solid loadings has been a target in lignocellulosic ethanol process designs. This is mainly to minimize costs directed related to volume, and also to keep low energy requirements for distillation. Working at high solid loadings had been considered to imply reduction in enzymatic performance, requiring also special equipment. Humbird *et al.* (2010) studied the effect of solid loading on economics, finding an optimum in the range of 15 - 20% (w/w). Taking into account NRELs optimum solids loading and the solid loading used in Novozymes internal trials mentioned above, it is considered for the design to work at 20% total solids (TS).

Enzymes kinetics can vary greatly, depending on hydrolysis conditions but also on biomass type and composition, and pretreatment method. Retention times for enzymatic hydrolysis of pretreated lignocellulosic biomass are typically in the 24 - 84h frame. For instance, NREL 2011 design, takes 84h for hydrolysis split in 24h of continuous high solid hydrolysis and 60h batch hydrolysis (2% enzyme loading). In contrast, NREL 2002 design assumes 72h of simultaneous saccharification and fermentation and the Brazilian CGEE takes 48h of separate saccharification (Aden *et al.*, 2002; Centro de Gestao e Estudos Estratégicos, 2009; Humbird *et al.*, 2011). On the other hand, it has been published that the change to newer enzyme preparations like Cellic CTec3 has resulted in savings related to lower dosing and shorter retention time. Thus, it is assumed that a hydrolysis retention time of 48h is required for 90% of glucan conversion at 4% enzyme loadings hydrolysis.

Enzymatic hydrolysis equipment type depends greatly on the material to hydrolyse. For high solids mixtures (> 15 %TS), there are several equipment examples at industrial scale, particularly in the anaerobic digestion of municipal solid waste (MSW). After a review of digestion systems for solid residues, see Appendix 7, parallel vertical plug-flow reactors are considered for the process. The reactors can be vertical towers in which the biomass flows downwards as it is continuously hydrolysed, similar to those described in 2002 NREL report (Aden *et al.*, 2002).

Table 27. Enzymatic reactor specifications

Specification	Units	R203
Inflowing streams	-	< 35 >
Outflow streams	-	< 36 >
Flow	tonne/h	382
Equipment Volume	m3	3500
Units required	-	6

7.1.3. *A300 – Waste Water Treatment*

Waste water streams, including POME and the condensed vent stream from flashed the pretreated slurry, are treated primarily in an anaerobic digestion system. It is considered that POME derived from a single mill is to be treated, assuming the design is integrated to a typical palm oil mill. The assumed composition of POME is presented in Table 13. POME is typically characterised by its COD content, and in order to simulate the conversion of COD to CH_4 and $CO_{2,}$ the incoming COD is attributed to oil and other organic material as $C_6H_{12}O_6$ (extractives component in the ASPEN model). Considering the average oil content in POME is 6 g/L, it is estimated that oil content contributes to 20.58 gCOD/L[4] to the total 50 g/L, the rest is allocated to other compounds, modelled in ASEPN as extractives with the formula $C_6H_{12}O_6$. Reactions in Table 28 are considered to take place in the anaerobic digestion reactor. Considering the composition of incoming POME, the achievable methane conversion is 0.234 kg CH_4/kg COD. Based on literature data of POME treatment, it is considered that 95% of COD in POME is removed with 15 days of hydraulic retention time (Yacob *et al.*, 2006). As other waste water streams are also treated, the conversions of residual carbohydrates and other components are presented in Table 28, and are also considered to be 95%. Reported values on nitrogen content in POME are simulated in ASPEN as ammonium acetate. Sludge formation is minimal and is simulated based on nitrogen in POME, assuming 95% of ammonium acetate into biomass. However, more detailed nitrogen balance and conversions are disregarded during the waste water treatment.

Table 28. Reactions considered in the Anaerobic waste water treatment

Reaction	Conversion
$C_{16}H_{32}O_2 + 7\ H_2O \longrightarrow 11.5\ CH_4 + 4.5\ CO_2$	95%
$2\ C_5H_{10}O_5 \longrightarrow 5\ CH_4 + 5\ CO_2$	95%
$C_6H_{12}O_6 \longrightarrow 3\ CH_4 + 3\ CO_2$	95%
$C_5H_4O_2 + 3\ H_2O \longrightarrow 2.5\ CH_4 + 2.5\ CO_2$	95%
$C_2H_4O_2 \longrightarrow CH_4 + CO_2$	95%

Table 29. Reactions considered in the aerobic stabilization of treated water

Reaction	Conversion
$C_{16}H_{32}O_2 + 23\ O_2 \longrightarrow 16\ H_2O + 16\ CO_2$	74%
$C_6H_{12}O_6 + 6\ O_2 \longrightarrow 6\ H_2O + 6\ CO_2$	74%
$C_5H_4O_2 + 5\ O_2 \longrightarrow 2\ H_2O + 2\ CO_2$	74%
$C_2H_4O_2 + 2\ O_2 \longrightarrow 2\ H_2O + 2\ CO_2$	74%

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[4] Palmitic acid has 3.43 gCOD/g acid and is used for the calculation of oil contribution to COD content in POME.

Following anaerobic treatment, it is considered that the waste water is stabilized. Aerobic conversion of remaining COD is considered for this purpose. Oxygen from the environment is used for oxidizing organics into CO_2 that is liberated in the atmosphere. Sludge is generated in this step of the treatment, it is separated and fed into the CHP system. Organics conversion into CO_2 is considered as 74% following NREL 2011 model on aerobic treatment (Table 29); 22% is considered for sludge formation. For treating large amounts of waste water, anaerobic and aerobic concrete digesters are assumed following NREL 2011 model (Humbird *et al.*, 2011). A summary of design assumptions and specifications are presented as Table 30.

Table 30. Specifications considered for the waste water treatment equipment

Specification	Units	R301	R302
Inflowing streams	-	< 47 > < 28 >	< 48 > < 50 >
Outflow streams	-	< 48 > < 49 >	< 21 > < 52 >
Flow	tonne/h	90	88
Residence time	d	15	-
Organics conversion	%	95	96
Temperature	°C	40	25

7.1.4. *A400 - Heat and Power*

The energy available in the solid residues and the recovered biogas is intended to cover the energy demands of the adjacent palm-oil mill and the design process, with any surplus ready for export. As both thermal and electrical energy are required, a combined heat and power system is necessary. Different technologies are available for generation of heat and power, each with its own advantages and disadvantages, see Appendix 7 – 1.6. After reviewing the available technology, it becomes clear that steam turbine systems are the best option.

Solid waste streams resulting from the process are fed into a furnace. Additionally, biogas is also considered as an energy source. The combustion reactions assumed in the furnace are presented in table 31. The heat is used to raised steam from preheated boiler feed water at 95 °C. It is considered that the system raises superheated steam to 454 °C and 63 bar. Boiler efficiency is taken as 80% and energy from flue gases is considered to be used in the system for preheating the fuel streams.

Table 31. Reactions considered for the simulation of solid residues combustion

Reaction	Heat of Reaction (J/mol)
$C_5H_{10}O_5 + 6\ O_2 \rightarrow 5\ H_2O + 5\ CO_2$	-2 269 120
$C_6H_{12}O_6 + 6\ O_2 \rightarrow 6\ H_2O + 6\ CO_2$	-2 722 944
$C_5H_4O_2 + 5\ O_2 \rightarrow 2\ H_2O + 5\ CO_2$	-2 300 178
$C_2H_4O_2 + 5\ O_2 \rightarrow 2\ H_2O + 2\ CO_2$	-837 848
$H_2SO_4 \rightarrow H_2O + 0.5\ O_2 + SO_4$	196 546
$CO_2 \rightarrow 0.5\ O_2 + CO$	282 980
Lignin $+ 8.5\ O_2 \rightarrow 4\ H_2O + 8\ CO_2$	-4 115 336
$2\ O_2 + CH_4 \rightarrow 2\ H_2O + 2\ CO_2$	-802 618
$6\ O_2 +$ Cellulose $\rightarrow 5\ H_2O + 6\ CO_2$	-2 593 768
$5\ O_2 +$ Hemicellulose $\rightarrow 4\ H_2O + 5\ CO_2$	-2 172 390
$1.3\ O_2 +$ Enzyme $\rightarrow 0.8\ H_2O + 0.24\ NO + CO_2 + .01SO_2$	-492 117
$Ca(OH)_2 \rightarrow H_2O + CaO$	109 182
$23\ O_2 + C_{16}H_{32}O_2 \rightarrow 16\ H_2O + 16\ CO_2$	-944 218
$1.33\ O_2 +$ Biomass $\rightarrow 0.82\ H_2O + 0.23\ NO + CO_2$	-473 906

Initially the process was defined and simulated without any recovery of energy amongst process streams. The energy requirements and target temperatures were identified, see Table 32. For the cooling of the pretreatment vent stream the enthalpy curve was analysed. It was decided to split the integration of the stream as two streams: 1) VENTa for condensing the stream at 99.5 °C and 2) VENTb for cooling it down to release temperature. On the other hand, the heat treatment of the sugars product stream is not integrated with other streams to prevent possible contamination.

From the duty information it was decided to integrate the preheating of the boiler feed water (A400 - BFWAT) with the cooling of the pretreated liquid (A200 – PTTLIQ) and POME. The heat exchanger network involving the streams in Table 32 consists then of two process heat exchanges and one condenser for the vent stream. The boiler feed water target temperature was initially set with a dT of 5 °C to the maximum temperature of available process streams. The model was programmed to calculate the amount of steam that could be raised from water at 95 °C and the heat available from combustion. As there is heat available from other streams to reach this temperature, it was considered a suitable preheating temperature. POME < 45 > is used to pre heat fresh boiler feed water < 55 > from 25 °C to 38 °C. Then, preheated BFW < 56 > is heated up to 95 °C with PTTLIQ < 23 >, as the later cools down to 45 °C. However, there is a small surplus of heat of 295 kW from both streams and cooling is required.

Table 32. Duty of hot and cold streams in the process

Stream		Initial Temp.	Target Temp.	Q
Type	Area-Stream name	°C	°C	kW
Hot	A200 - PTTLIQ	100	45	-8 848
Hot	A200 - VENT1a	100	99.5	-30 986
Hot	A200 - VENT1b	99.5	40	-3 413
Hot	A300 - POME	80	40	-1 808
Cold	A400 – BFWAT	25	95	10 361

7.2. Ten-mill case

The ten-mill case process and equipment design is based on the same consideration as for the three-mill case. It is considered that the availability of solid biomass residues is increased from 3 to 10 mills processing 60 tonne/h of FFB each.

7.2.1. *A100 – Pre-processing*

The pre-processing equipment for mechanical manipulation of the incoming biomass is designed following the same considerations as in section 7.1 – Three-mill Case. The difference for achieving higher processing capacity relies in the amount of units, or tandems in the case of milling operations. See Tables 33 and 34 for considered specifications.

Table 33. Specifications for pre-processing equipment

Speficication	Unit	A101	A103	A105
Incoming streams	-	< 5 >	< 2 >	< 10 >
Outflowing streams	-	< 6 >	< 4 >	< 11 >
Processing capacity	tonne/h	367	261	138
Power consumption	kW/dry tonne	18	20	18
Power	kW	1925	1566	775

Table 34. Milling tandem specifications

Specification			Unit	A104
Input	PBM	< 7 >	tonne/h	628
	Water	< 14 >	tonne/h	85
Output:	Sap	< 9 >	tonne/h	595
	PBMF	< 8 >	tonne/h	315
Fibre fraction			w/w	0.3
Roller speed			rpm	6
Number of rollers (incl. crusher)			-	14
Tandems			-	12
Power consumption			kW/tph	22
Total power			kW	13813

7.2.1. *A200 – Biomass Conversion*

Equipment characteristics in biomass conversion area were estimated following the same assumptions as in the three-mill case. The main characteristics necessary for quoting the equipment are presented in tables 35 through 39.

Table 35. Pretreatment system specifications

Specification	Units	E201	R201
Incoming streams	-	< 13 > < 68 >	< 15 > < 18 > < 62 >
Outflow streams	-	< 15 >	< 20 >
Flow	tonne/h	835	302
Residence time	min	10	2
Design volume	m3	123	45
Temperature	°C	100	190

Table 36. Specifications for screw press equipment

Specification	Units	S202	S204
Incoming steams	-	< 21 >	< 36 >
Outflow streams	-	< 23 > < 24 >	< 37 > < 38 >
Flow	tonne/h	845	1274
Max. Capacity	tonne/h	140	140
Units	-	7	10
Av capacity used	tonne/h	121	127
Power per unit,	kW/unit	108	109
Total power	kW	1296	1955

Table 37. Conditioning reactor specifications

Specification	Units	R202
Inflowing streams	-	< 25 >
		< 28 >
Outflow streams	-	< 29 >
Flow	tonne/h	504
Residence time	min	20
Design volume	m3	225
Agitator tip speed	m/s	3.3
Agitation Power	kW	22

Table 38. Rotary drum filter specifications

Specification	Unit	S203
Inflowing streams	-	< 29 >
Outflow streams	-	< 30 > < 31 >
Flow	tonne/h	530
Units required	-	9
Total Power	kW	1311

Table 39. Enzymatic hydrolysis reactor specifications

Specification	Units	R203
Inflowing streams	-	< 35 >
Outflow streams	-	< 36 >
Flow	tonne/h	1172
Equipment Volume	M3	3516
Units required	-	16

7.2.1. *A300 – Waste Water Treatment*

Similarly to the other process areas, equipment characteristics in waste water treatment area were estimated following the same assumptions as in the three-mill case. The main characteristics necessary for quoting the equipment are presented in Table 40.

Table 40. Waste water treatment equipment

Specification	Units	R301	R302
Inflowing streams	-	< 47 > < 28 >	< 48 > < 50 >
Outflow streams	-	< 48 > < 49 >	< 21 > < 52 >
Flow	tonne/h	206	203
Residence time	d	15	-
Organics conversion	%	95	96
Temperature	°C	40	25

7.2.1. *A400 - Heat and Power*

Similarly to the approach used for the three mill case, the process streams were first analysed before any heat recovery, Table 41. The target temperature for feed boiler water was initially set as 95 °C as in the previous case, which is an input for estimating the amount of steam that can be raised, see Table 41. It is clear that by integrating the same streams as in the three-mill case, the cooling requirement is less than the requirement for preheating the boiler feed water to 95 °C. Considering these findings, the boiler feed water is preheated only up to 92 °C. As a result, the proportion of raised steam per amount of processed biomass is slightly lower than in the three mill case.

Table 41. Duty requirements of hot and cold streams in the process

Stream		T	T end	Flow	Q
TYPE	Area-Name	°C	C	tonne/h	kW
Hot	A200 - PTTLIQ	100	45	492	-29 624
Hot	A200 - VENT1a	100	99.5	165	-103 456
Hot	A200 - VENT1b	99.5	40	165	-11 393
Hot	A300 - POME	80	40	40	-1 808
Cold	A400 – BFWAT	25	95	415	33 110

8. Economic Evaluation

8.1. Raw Materials Cost

8.1.1. *Biomass Cost*

According to the National Biomass Strategy published by the Agensi Inovasi Malaysia (2011), the price allocation for the mentioned biomass must consider four different cost: (1) substitution, (2) harvesting plus collection, (3) pre-processing, and (4) transportation. Substitution cost is considered as replacement cost for EFB, OPT, and OPF, whereas for shells and fibres it is based on market price. Harvesting and transportation costs are not considered for EFB, fibres, and shells, as they are readily available at the mill. Pre-processing cost, mainly related to drying as mentioned in the National Biomass Strategy report, are not accounted in the calculations as this would be inside the battery limits. The list of costs is presented in Table 42, taking the average of values given in the mentioned strategy report[5]. Transportation cost of all biomass residues is based on a plantation area of 18 000 ha per mill and considering an average distance, as described in Section 3.2 – Basic Assumptions. The costs presented in Table 42 apply for Cases A and B in Section 2.5 - Process Alternatives.

Table 42. Biomass costs for single mill related biomass, in RM/tonne wet-basis

Biomass type	Substitution	Harvesting*	Transport	Total
EFB	93	0	0	93
OPF	35	12	26	73
OPT	78	14	26	118
Shell	130	0	0	130
Fibre	40	0	0	40

* Fresh OPF and OPT contain 70% moisture.

In Figure 16, the change in biomass cost, given in RM per tonne of available biomass, is presented as a function of the number of mills (thus transportation area). It is clear that for lower scales, mill biomass results cheaper as it is already available at the processing (central) mill, thus no transportation cost is allocated to this fraction of the total biomass.

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[5] All prices are reported on wet basis except for harvesting cost. In the presented table it is taken that 1 tonne of OPT (or OPF) contains 0.7 tonnes of dry weight.

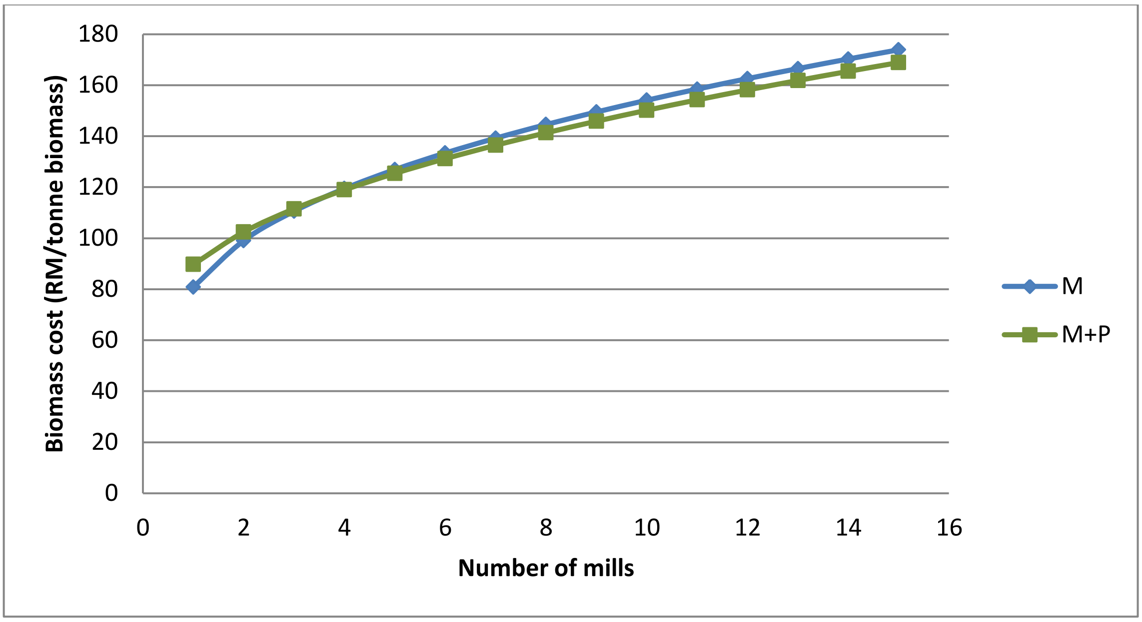

Figure 16. Cost of biomass and number of mills

M: Mill biomass - Empty fruit bunches, shells and fibres. M+P: Mill and Plantation biomass - oil palm fronds and oil palm trunks.

Thus, in the present design the cost of biomass considered as input for the three- and ten-mill cases is given in Table 43. The total input of biomass as available to the central mill (OPT, OPF, EFB and Fibres) results in costs of 110 RM/tonne and 148 RM/tonne to the three- and ten-mill cases respectively.

Table 43. Biomass cost difference as result of transport, all in RM/tonne

	Three-mill Case	Ten-mill Case
EFB	123	166
OPF	92	128
OPT	137	173
Fibre	70	113

8.1.2. *Utilities and other streams*

The utilities prices, presented in Table 44 are in accordance to the Malaysian Investment Development Authority (2012). Steam price is taken from SuperPro Designer's databank, on energy content basis (Inteligen). Although part of the steam is intended for export to the adjacent mill, as the biomass derived from the same mill carries a cost, the analysis also includes a cost to exported steam. Gypsum disposal costs are based on reported fees paid by industries delivering solid waste to Krubong landfill (Hun Yang *et al.*, 2009). The cost of lime and acid are taken from the Institution of Chemical Engineers price listing; although it is intended for educational purposes is still in accordance with current prices found on the market (Pitt, 2002).

Table 44. Cost and prices of various inputs and outputs

Utility	Cost
Electricity (RM/MWh)	283
Enzymes[6] (RM/tonne)	4128
Fermentable sugars (RM/kg)	1.77
Gypsum (RM/5 tonne load)	8
Lime (RM/tonne)	684
Acid (RM/tonne)	349
Steam (RM/MWh)	66.2

When necessary, conversion from Malaysian Ringgit to US dollars and to euros are taken 0.324 USD/RM and 0.25 EUR/RM based on the average conversion rates in the period from November 2011 to May 2012. These conversion rates are used throughout the whole report, historical prices are updated as mentioned in section 8.3 – Total Capital Investment.

8.2. Fermentable Sugars Price and Market

The availability of renewable feedstock has been identified as a key aspect to incentivize the bio-based chemical industry growth in Malaysia. Moreover, fermentable sugars have been recognised as the largest feedstock that can supply the growing bio-based chemical industry (National Research Council, 2000). This is supported by the fact that sugars are the initial component of plenty of well-established fermentations for production of commodities and fine chemicals (such as citric and ascorbic acids) and holds a large potential in emerging bio-based products. In this context, the production of fermentable sugars derived from already available palm oil solid waste is an appealing alternative for integration at mill facilities. Thus, the use of second-generation sugars would support Malaysia in the global transition towards a bio-based economy.

Although fermentable sugars hold a large economic potential in a bio-based economy, a fermentable sugar product, as an intermediate, is not yet found on the market. For this economic evaluation, molasses (a sugar by-product sometimes used for industrial fermentations) and unrefined raw sugar are taken as an indication of the fermentable sugar market. In Malaysia, the sugar industry imports most of its raw material; the largest sugar refiner imports 97% of its raw sugar from Brazil and Australia (CIMB Reseach, 2011). Sugar prices have been historically regulated in Malaysia, and in early 2012 a new long-term contract (LTC) was signed for raw sugar supply into the country. The new

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[6] Calculated from 0.12 US dollars per gallon of ethanol produced, was calculated back to sugar assuming 0.511 kgethanol/kgsugar, (Humbird, et al. 2011).

contract fixes the raw sugar supply price at 1.77 RM/kg until 2014, with the objective to mitigate the volatility of sugar markets (Ariffin, 2012).

Taking the LTC sugar price as a base, an economic evaluation is performed following the proposed design for both three- and ten-mill cases. The profitability of the project is estimated from production cost estimations and revenues, taking the LTC sugar price for the fermentable sugars product. The two cases, evaluated throughout the report, differ thus in capital expenditure as the ten-mill case is benefited by the economies of scale. Thus, operational expenditures show benefits related to economies of scale (labour and capital charges), but also show higher raw material expenditures with scale, as transportation cost per tonne input biomass increases.

As the profitability of the two cases considered through this report is related to capital investment and biomass (or its transportation) costs, sensitivity analysis in which these parameters are varied are performed. Furthermore, considering the market volatility and the relative short duration of the LTC, sugar price is varied approaching the lowest and highest raw sugar prices in the international markets in the past three years, as mentioned in section 2.6.

8.3. Total Capital Investment

Total Capital Investment (TCI) is calculated based on individual costs for main equipment. The individual equipment cost is listed in Appendix 7. Historical equipment cost are updated using US inflation indices, e.g. equipment cost from 2002 are updated to 2011 USD values. Installation, piping and instrumentation are accounted as multipliers to the equipment purchase cost individually, according to reported values to each equipment. The obtained value comprising direct and indirect costs for each equipment are accounted as total bare module cost, C_{TBM}. From the sum of the total bare module cost, the total capital investment is calculated following equation 5, based on Guthrie method (Seider *et al.*, 2010).

$$C_{TCI} := 1.18 \cdot \left(C_{TBM} + C_{site} + C_{buildings} + C_{offsite}\right) + C_{wc} \quad \text{(equation 5)}$$

Where:

C_{TCI} = Total Capital Investment,
C_{TBM} = Total Bare Module Cost,
C_{site} = Site Development Cost,
$C_{buildings}$ = Building Cost,
$C_{offsite}$ = Offsite Facilities cost,

C_w = Working Capital.

Site development cost can be estimated as 10 – 20% of C_{TBM} for grass roots plants, or as 4 - 6% for integrated complexes. Considering that the proposed design is to be integrated in already existing facilities, the site development cost are taken as 4% for the three-mill case and 6% for the ten mill case as the impact of a larger facility to the already existing mill site is considered greater. In the case of building cost, it is taken as 15% of C_{TBM} accounting for process (10%) and non-process buildings (5%) as typically estimated for projects to be integrated to existing complexes (Seider *et al.*, 2010). Offsite facilities cost includes utility plants, pollution control, ponds, waste treatment, and other possible facilities that had not been accounted for in the modules. In this case, waste water treatment and power and heat production are already accounted in C_{TBM}, thus the only contribution to the offsite facilities cost factors is based on cooling requirements. Individual components of the TCI are given in Table 45 for the three- and ten- mill cases. The scale index describing the difference of capital investment amongst both cases is 0.63, lower than the 0.72 index previously estimated in Section 2 - Process Definition. The lower index is due to the large impact the boiler/turbine system have in the total equipment cost (Figure 17), both with scale indices of 0.6, and which were not accounted for in detail in the estimations given in Section 2 - Process Definition.

Table 45. Capital Investment, three-mill case

Investment Contribution		Three-mill Case		Ten-mill Case	
		USD x 10^3	RM x 10^3	USD x 10^3	RM x 10^3
Total Purchased Equipment Cost	C_{TPE}	66 852	206 332	138 053	426 089
Total Bare Module Cost	C_{TBM}	105 157	324 559	221 919	684 936
Site Development Cost	C_{site}	4 206	12 982	13 315	41 096
Buildings Cost	$C_{building}$	15 774	48 684	33 288	102 740
Offsite Facilities Cost	$C_{offsite}$	52	162	126	390
Total Permanent Investment	C_{TPI}	147 724	455 937	317 005	978 411
Working Capital Cost	C_{WC}	25 999	80 245	55 793	172 200
Total Capital Investment		**173 723**	**536 182**	**372 798**	**1 150 612**

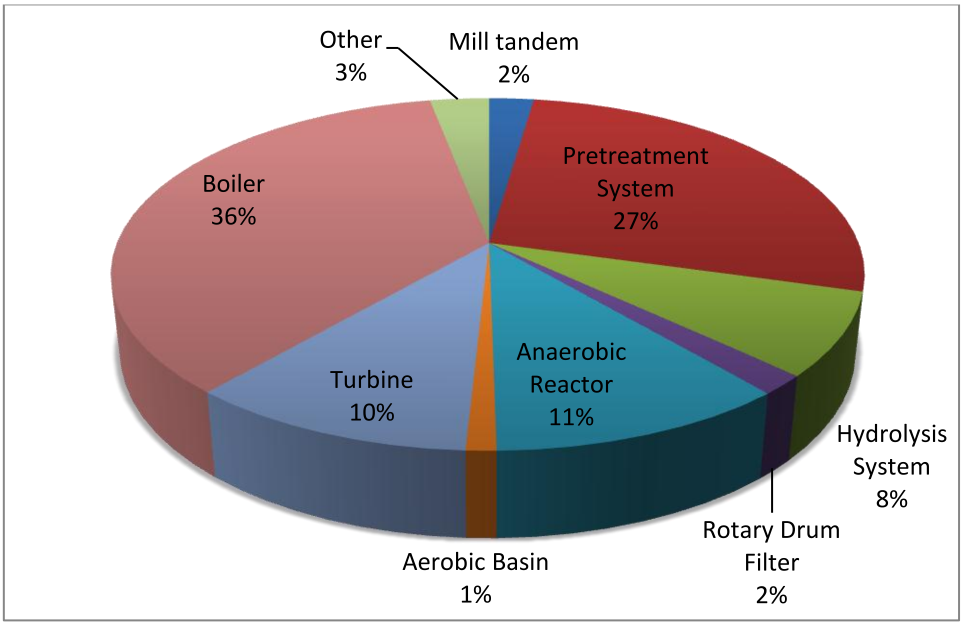

Figure 17. Equipment contribution to total equipment cost

8.4. Total Production Cost

Total production cost (TPC) is calculated as a sum of direct production costs (DPC), fixed charges, plant overhead, and general expenses. The main contributions to DPC are raw materials (biomass and other raw materials such as enzymes), labour, and maintenance. Fixed charges or capital charge, are taken as constant charges for considering depreciation and interests on the fixed capital throughout the project life, equation 6.

$$C_{cap} := \frac{i}{1-(1+i)^{-n}} \cdot C_{TDC} \qquad \text{(equation 6)}$$

where:

C_{TDC} = Depreciable Capital (TCI excl. working capital),

C_{cap} = Annualized capital charge,

i = Interest rate in Malaysia taken as 6.5%, based on the highest lending rate reported in 2011 and that in the first quarter of 2012 (Central Bank of Malaysia, 2012).

Biomass cost is calculated as described in 9.2.1 – Biomass Cost, where transportation cost are already taken into account. Labour and other DPC costs are calculated as function of C_{TPI}, number of operators with a base wage cost of 90 RM/d (CIDB, 2012), taking typical considerations for solid-fluids handling processes. Number of operators was estimated for the three mill case and escalated to the ten-mill case assuming a 0.25 exponent for the scale ratio, as recommended by Peters *et al.* (2003). Plant overhead expenses, are calculated in relation to labour and maintenance (Seider *et al.*, 2010). General expenses, including administrative and sales charges, remain constant per tonne of produced sugar as they are calculated in relation to sales (Seider *et al.*, 2010). Integration of the process into milling facilities implies reduction of certain overhead costs, and are accounted in the estimations. The main contributions considered in TPC calculations are given in table 46 and 47, for more detailed considerations see Appendix 7 – Economic Evaluation.

Table 46. Production cost components, in RM x 10^3 per year

Cost Contribution		Three-mill Case	Ten-mill Case
Capital Charge	C_{cap}	63 423	136 102
Biomass (incl. transportation cost)	C_{BM}	174 839	787 483
Raw Materials	C_{RM}	8 507	29 054
Waste Disposal	C_{waste}	19	67
Total Labour Cost	C_{lab}	3 600	4 874
Total Maintenance Cost	C_{maint}	46 164	101 266
Overhead	C_{overh}	4 533	13 561
Insurance and Taxes	C_{tax}	9 119	19 568
General Expenses	C_{gral}	73 293	245 086
TOTAL PRODUCTION COST	TPC	383 497	1 337 061

Table 47. Total production cost comparison of three-mill and ten-mill cases

Total Production cost		Three-mill Case	Ten-mill Case
Annual Production Cost	USD/y x 10^3	124 523	433 208
	RM/y x 10^3	388 497	1 337 061
Specific Production Cost*	USD/tonne	348	363
	RM/tonne	1074	1120

*Based on the dry weight of produced sugars.

As already estimated in the Section 2 - Process Definition, the transportation cost affecting the production cost have a greater impact than the economies of scale represented in the capital charge fraction of the total production cost. The difference in specific production cost amongst both cases is 47 RM/tonne of sugar (*ca.* 5%), slightly lower than estimated in Section 2 due to the lower overall scale index describing the total capital investment change with scale.

8.5. Revenue

A potential revenue for both cases is calculated considering the LTC price for sugar as reference and the current 25% corporative tax in Malaysia (Malaysian Investment Development Authority, 2012), Table 48. Total sales account for the fermentable sugars product and other co-products, while straight-line depreciation is considered for ten years.

Table 48. Revenue comparison of three-mill and ten-mill cases

	Three-mill Case		Ten-mill Case	
	RM/y x 10^3	RM/tonne*	RM/y x 10^3	RM/tonne*
Total Sales	683 644	1 914	2 281 857	1 911
Operating Income	363 570	1 003	1 080 898	905
Depreciation	45 594	128	97 841	82
Earnings Before Tax	317 976	876	983 057	823
Earnings After Tax	238 482	657	737 293	617

*Based on the dry weight of produced sugars.

8.6. Profitability Analysis

In order to evaluate the profitability of the project, it is assumed that the plant is constructed at the end of two years. The TPI are accounted as cash outflows during the construction of the plant. Thus, TPI cost are split in the two first years, while the working capital for start-up is discounted only on the 2nd year. The first operational year is the 3rd year of the project, and constant cash flows are assumed throughout ten years of operational project. All the cash out- and inflows (CF) are converted to present values, as described in Appendix 7 – Economic Evaluation. Considering the proposed design project, at the given costs and prices and under the current situation in Malaysia with 6.5% interest rate and 25% taxes, the net present value (NPV) of the three-mill project is 1 315 million RM with an internal rate of return (IRR) equal to 44% and a payback after the 3rd year of operation (PBT), Figure 18. In contrast, the larger ten-mill project has an NPV of 4 250 million RM with an IRR of 57%, Table 49. If the net present value of both cases is calculated per tonne of sugar produced in the project life, the three mill case value is 368 RM/kg sugar while the ten-mill case value is 356 RM/kg of sugar.

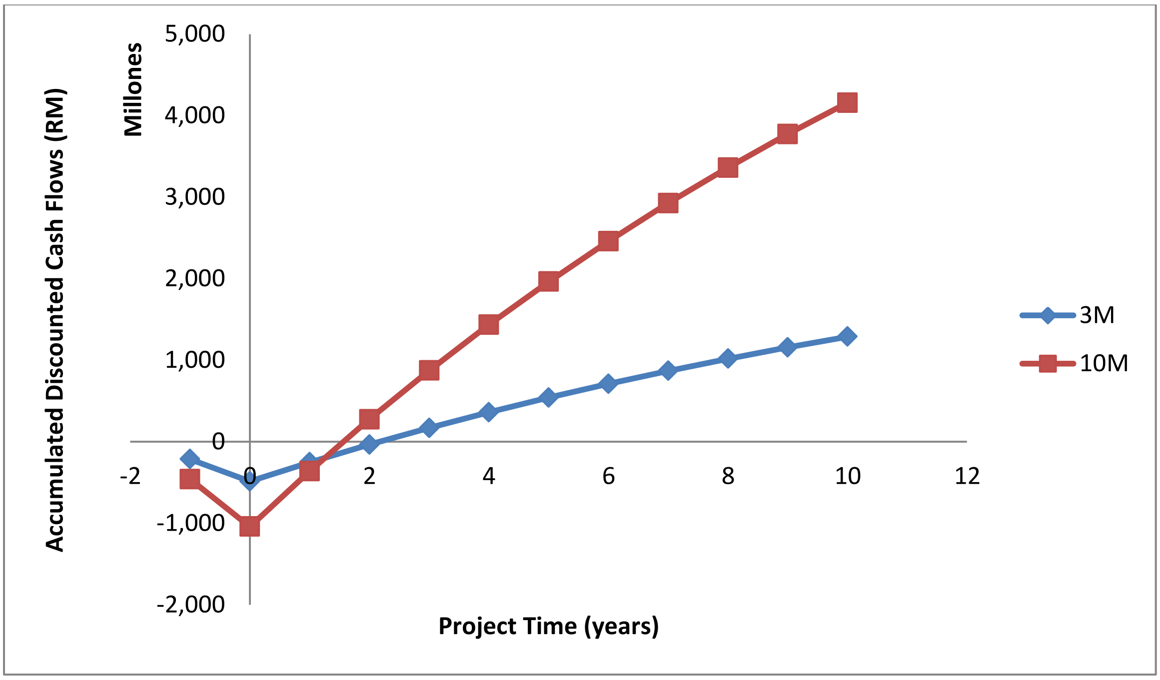

Figure 18. Discounted cash flows for, at 6.5% interest rate and 10 years of operation
3M – Three-mill case; 10M – Ten-mill case

Table 49. Comparison of profitability analysis

	Three-mill Case	Ten-mill case
Break-even year*	3rd	2nd
IRR (%)	44	57
NPV (million RM)	1 315	4 250
Specific NPV (RM/kg sugar)**	368	356

*Year after start-up.

**NPV divided by tonnes of sugar produced in the project life.

8.7. Sensitivity Analyses

The main factors that impact differently the economic evaluation of the project as scale changes are biomass cost (transport distance) and the capital investment. Thus, these two parameters are varied in order to show the impact these have on the economic potential of the project, Additionally, due to the uncertainty of fermentable sugar price a sensitivity analysis considering possible ranges for this factor is presented. Finally, the interest rate assumed for the project is changed to 10% (+ 15.4% of the Malaysian base lending interest rate), which is a value typically used in economic analysis.

8.7.1. *Transportation distance*

As the transportation cost have been identified as one of the factors affecting the profitability of the process in relation to scale, this parameter is varied to see its effect in both scales. As the transportation is based on plantation area and reported yields, smaller distances are considered unrealistic. Instead, larger distances which could be due to plantations and mills separated from each other, are considered. Thus, a sensitivity analysis in which the average distance is increased is presented in Table 50. It is clear that higher distances have higher impact on the profitability of the ten-mill case than in the three-mill case. In the ten-mill case, TPC cost equals the LTC sugar price (thus gross earnings are zero), when the distance is increased to about 1.8 times the average distance when the mills and plantations are adjacent to each other. In contrast, this TPC value is achieved when the average distance is about 4 times the average distance when the mills and plantations were adjacent to each other.

Table 50. Sensitivity Analysis on transportation distance, all in RM/tonne of sugar*

Distance increase (%)	Three-mill Case		Ten-mill Case	
	Specific TPC	Specific NPV	Specific TPC	Specific NPV
0	1.07	368	1.12	356
15	1.10	356	1.17	332
30	1.13	343	1.23	308
50	1.17	327	1.30	276
100	1.25	288	1.47	196
180	1.41	215	1.76	67

* NPV values (RM) are divided by tonnes of sugar produced in the project life, TPC values (RM/y) are divided by the yearly sugar production.

8.7.2. *Price of sugar*

Already in section 9.1 the uncertainty of a fermentable-sugar product price was mentioned. Although the LTC price fixes the price of raw sugar in Malaysia, this does not necessarily applies to fermentable sugars that can be used in the bio-based industry. Looking at historical sugar prices worldwide in the past years, a sensitivity analysis reducing the sugar price 15 and 30% is performed; the rest of the estimated cost are kept the same as in the original estimation. The results of this sensitivity analysis are given in Table 51, where it can be seen that changes in the price of sugar have a large impact on the profitability of the project, but the difference amongst both scales remains the same.

Table 51. Sensitivity Analysis on Sugar Product Price

	Three-mill Case			Ten-mill Case		
	- 30%	- 15%	LTC	- 30%	- 15%	LTC
Sugar Price (RM/kg)	1.24	1.51	1.77	1.24	1.51	1.77
Break-even year*	5th	3rd	3rd	4th	3rd	2nd
IRR (%)	20	33	44	24	42	57
NPV (million RM)	410	861	1 315	1 225	2 737	4 250
Specific NPV (RM/kg sugar)**	115	241	368	103	229	356

*Year after start-up.

**NPV divided by tonnes of sugar produced in the project life.

8.7.3. *Total Capital Investment*

Considering the rough equipment cost calculations, the impact of varying the total TCI +/- 30% is analysed. By varying this factor, also the total production cost changes as the capital charge and maintenance cost are in function of the capital investment. Compared to the observed changes in the sugar product price sensitivity analysis however, it can be seen that TCI possible changes have less impact on the profitability of the project.

Table 52. Sensitivity Analysis results on total capital investment

	Three-mill Case			Ten-mill Case		
	+ 30%	- 30%	TCI	+ 30%	- 30%	TCI
TCI (million RM)	697	375	536	1 495	805	1 150
Break-even year*	3rd	2nd	3rd	3rd	2st	2nd
IRR (%)	32	62	44	44	79	57
NPV (million RM)	1 106	1 523	1 315	3 794	4 707	4 250
Specific NPV (RM/kg sugar)**	310	427	368	317	394	356

*Year after start-up.

**NPV divided by tonnes of sugar produced in the project life.

8.7.1. *Interest Rate*

The interest rate assumed for the project is changed to 10% (+ 15.4% of the Malaysian base lending interest rate), which is a value sometimes used in economic analysis. Due to the larger capital investment, it is clear that a higher interest rate has a higher impact on the profitability of the ten-mill case. However, the profitability of the project remains similar in both cases, when compared to the sensitivity of the project's profitability to the sugar product price, Table 53.

Table 53. Sensitivity analysis results on interest rate

	Three-mill Case		Ten-mill Case	
	10%	6.5%	10%	6.5%
Break-even year*	3rd	3rd	2nd	2nd
NPV (million RM)	981	1 315	3 249	4 250
Specific NPV (RM/kg sugar)**	275	368	272	356

*Year after start-up.

**NPV divided by tonnes of sugar produced in the project life.

8.8. Evaluation

Following the profitability analysis presented above, it is clear that the project is profitable in both scales, given the price and transportation distances assumed herein. As larger scales imply higher selling volumes, the project at ten-mill scale case has higher NPV. However, when the amount of fermentable sugars produced in the project life is considered, the present value of the project becomes slightly higher for the three-mill case, with 368 RM/tonne of sugar produced *vs.* 356 RM/tonne sugar for the ten-mill case.

Additionally, the sensitivity analysis show that the three-mill case is more robust to changes on transportation distance, which is highly relevant considering that mills from a same company are not always adjacent to each other. Changes in sugar prices clearly affect the project at any scale, however, given the higher production cost (incl. capital charges) given at larger scales, the profitability of the ten-mill case is affected more drastically surpassing benefits related to higher selling quantities.

Overall, the project is profitable in both cases, with payback in the third year of operation and similar total production cost; 1 074 and 1 120 RM/tonne sugar for the three- and ten-mill cases respectively. Considering that the risks are higher for the ten-mill case, and that the profitability of the project is more sensible to changes in transportation distance and sugar prices at higher scale, it is considered that the project at a three-mill case is a better alternative.

9. Life Cycle Analysis

Considering the environmental concerns related to palm oil in the last years, a life cycle assessment (LCA) of the proposed process is performed following as close as possible the guidelines presented as ISO 14044:2006. The goal of the assessment is a preliminary evaluation of the process for the production of second generation sugars, in order to have an initial estimate of the impact it can bring to palm oil production. This assessment is intended for the plantation, milling, refining and chemical partner companies to estimate the environmental benefit attainable with the proposed process. Knowing the impact of the proposed process is of relevance given the positive effects that waste conversion into second generation products can bring to the whole palm oil value chain, and can be later developed to evaluate it impact in the plastics or other chemicals obtainable from the second generation intermediate sugars.

9.1. Functional Unit

Although the final product from the proposed process is an intermediate for second generation chemical products, this LCA study is related to amount of produced palm oil instead, with a functional unit defined as one tonne of palm oil. The selection of this functional unit allows to have an initial estimation of the environmental benefits the integration of the proposed process can bring to the palm oil production chain.

9.2. Life Cycle Assessment Scope

As the proposed process is intended for its integration into the palm oil production chain, the boundaries of the system are expanded to include the palm oil plantation and milling operations. Furthermore, the system has also been expanded to include the production of crude palm kernel oil. Crude palm kernel oil has been identified as exchangeable with crude palm oil, so the proportional production of both oils related to the same amount of FFB is considered. In other words, the comparison of MICCI palm oil (palm oil production coupled with the proposed MICCI process) *vs.* typical palm oil, considers the effects related to the production one tonne of crude palm oil, from both fruits and kernels.

As can be seen in Figure 19, several economically valuable products are obtainable form the considered system, namely CPO, crude PKO, PKC and shells from palm oil milling operations, and a fermentable sugars (FS) Product. The resulting environmental impact is thus related to these products, as described below in the system delimitation section below.

Palm oil refinery operations are not considered in the life cycle, however the evaluation is based over the amount of CPO and crude PKO necessary to produce one tonne of palm oil. Thus, the evaluation is calculated over 0.93 tonnes of CPO and 0.11 tonnes of crude PKO, which are the main inputs necessary to produce one tonne of palm oil.

9.3. System delimitation

In order to make a fair distribution of the impacts associated to common operations related to these products, the system can be studied following different LCA approaches. The impact of the whole or parts of the process can either be distributed based on allocation methods, or the system itself can be expanded to separate products and operations. Following recommendations given in ISO 12044:2006, the present system is expanded, covering separately agricultural, milling and solid residues conversion. However due to limitations in information and/or added complexity, allocation in some parts of the system is still required.

The LCA model is constructed by defining four main processes, schematized in dotted lines in Figure 19. These processes are: (1) Plantation operations, (2) MICCI cluster operations, (3) PKO milling, and (4) Enzyme production. MICCI cluster operations imply the operations related to the three-mill case as discussed in the report Thus, the MICCI cluster operations covers activities from three separate palm oil mills as well as MICCI proposed design (with power and heat generation, and water treatment) adjacent to one of the three mills.

The overall balance over the MICCI cluster is derived considering that two typical mills deliver their biomass residues to a solid-biomass processing plant (MICCI Plant) that also has an adjacent third mill from which it also receives solid biomass. Additionally, the MICCI plant receives OPT and OPF from the plantations related to the three palm oil mentioned mills. POME treatment and energy requirements related to the adjacent mill is considered inside the proposed MICCI process. Energy from the MICCI process is considered for the energy demands of the whole MICCI cluster, with remaining requirements covered by other energy means as discusses by Schmidt (2007).

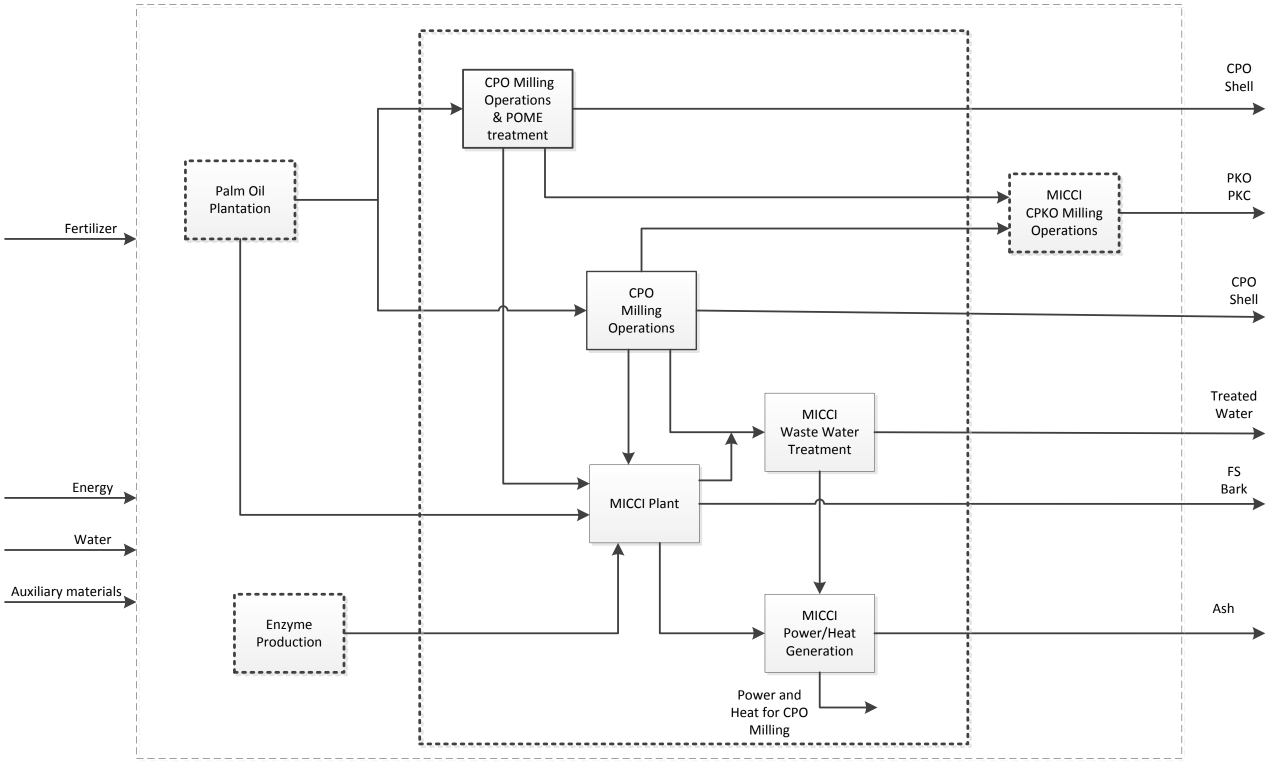

Figure 19. Life Cycle Assessment Boundaries

Slashed line – Life Cycle Assessment boundaries. Dotted lines – Main Processes boundaries. MICCI – Mill integrated conversion of commodity intermediaries (proposed design). CPO – Crude palm kernel oil. PKC – Palm kernel cake. PKO – Palm kernel oil. FS – Fermentable sugars. POME – Palm mill oil effluent.

Following the recommendations in ISO 14044:2006, allocations conflicts are avoided as much as possible. This approach was followed by considering co-products as displaced products in the techno-sphere. Therefore, it is considered that: (1) tree bark is replaced by the outgoing OPT bark stream, (2) sand is replaced by the outgoing ash stream, (3) hard coal for burning in power plant is replaced by shells and, (4) sugar production is replaced by the FS product stream. The remaining products are thus only CPO, PKO and PKC. The allocation method used for these remaining products is based on an economic basis; palm oil (from both CPO and crude PKO) with allocation of 98% and PKC with the remaining 2% (Schmidt, 2007).

Two different impact inventories for sugar can found in the accessed databases, which could be set as avoided interventions when considering FS product as sugar replacement. These inventories are related to sugar produced from sugarcane and from sugar beet. A comparison of setting the FS product as replacement for both types of sugars is made, taking into account the production chain related to the sugar, not the final end-use.

Additionally, in order to make a fair comparison of the proposed process in the cluster setting *vs.* typical milling operations, a separate assessment is made on typical palm oil milling. Thus, similar assumptions are considered, with inventories based on the same databases.

9.4. Life Cycle Inventory

In this evaluation, sub-processes delivering inputs to the proposed process were entered with inventory data reported by Schmidt (2007) on palm oil production in Malaysia and Indonesia. Inventories of oil palm plantation and palm oil milling activities are taken from the mentioned study, however the inputs and outputs related to POME treatment (in one of the three mills) and power generation were excluded as these activities are taken into account in the proposed MICCI proposed process. The overall MICCI process inventory is given in Appendix 8 together with inventories for palm kernel milling and oil palm plantation activities. The rest of the interventions related to the oil production were taken from ecoinvent and DK Food databases, as used by the same author. The only exception was the cellulose production since no cellulase process was defined in the accessed databases. The cellulase process was entered following the inventories given by U.S. Life Cycle Inventory Database (National Renewable Energy Laboratory, 2012), which is included in Appendix 8.

In order to avoid allocation of interventions, the system was extended beyond the proposed design, and material displacement was considered for co-product streams. This co-product definition implies that as result of implementing the proposed project certain activities and process are avoided, e.g. sugar extraction, and thus the environmental effects are positive. Due to the current study limitations, biogas recovered from POME treatment in the two peripheral palm oil mills is also defined as a displacing co-product instead of integrating its combustion in the milling process. This assumption is considered valid as the displaced product was defined as agricultural biogas production for intended use on-site, which is closely related to the project consideration.

9.5. Life Cycle Impact Assessment

The selected methodology for the life cycle impact assessment (LCIA) is EDIP2003. This methodology was selected as it covers several impact categories, such as global warming, ozone depletion, acidification, eco-toxicity, amongst others that are considered relevant. As a point of comparison conventional palm oil life cycle is evaluated. The network trees of the conventional and proposed processes can be found in Appendix 8.

The characterised results from the assessment are presented in Table 55. In most categories the integration of the MICCI process results in decreased impacts, whether the FS product is considered a replacement of sugar from sugar cane or sugar beet. The categories in which impacts are mostly decreased when compared to conventional oil production are human toxicity to air and soil, and ozone formation, most notoriously when it is considered that sugar from sugarcane is replaced.

Table 54 Life Cycle Impact of palm oil production (EDIP 2003 methodology).

Impact category	Unit	Conventional Oil	MICCI SC Oil	MICCI SB Oil
Acidification	m2	172.32	91.32	0.07
Aquatic eutrophication EP(N)	kg N	13.09	12.36	0.01
Aquatic eutrophication EP(P)	kg P	0.44	0.34	0.00
Bulk waste	kg	310.42	267.44	-0.01
Ecotoxicity soil chronic	m3	224 054.75	202 187.48	147.08
Ecotoxicity water acute	m3	68 070 877	67 992 069	64 195
Ecotoxicity water chronic	m3	1 859 275 993	1 858 077 947	1 753 863
Global warming 100a	kg CO2 eq	1 360	2 353	1.75
Hazardous waste	kg	0.01	0.01	0.00
Human toxicity air	m3	13 950 072	-317 983 049	-10 488
Human toxicity soil	m3	188.61	-646.23	0.02
Human toxicity water	m3	12 360	9 323	62.23
Ozone depletion	kg CFC11 eq	0.00	0.00	0.00
Ozone formation (Human)	person.ppm.h	0.17	-1.79	0.00
Ozone formation (Vegetation)	m2.ppm.h	2 404	-22 684	-2.93
Radioactive waste	kg	0.01	0.00	0.00
Resources (all)	kg	0.15	0.21	0.00
Slags/ashes	kg	0.28	-1.84	0.00
Terrestrial eutrophication	m2	584.65	340.58	0.22

MICCI SC – FS Product as sugar from sugarcane replacement; MICCI SB - FS Product as sugar from sugar beet replacement

In order to have an indication of the changes associated to all categories, the results are normalized and weighted following the same EDIP 2003 methodology. Thus, all the impact categories can be accounted on a single score basis and be compared to each other. It can be seen that the highest impact related to conventional palm oil is on the aquatic and terrestrial eutrophication, followed by human toxicity and terrestrial eutrophication, Figure 20.

In accordance with Table 54, after analysing the MICCI palm oil life cycle on a single score basis, it is clear that substantial benefits are incorporated, whether the FS product is set to replace either sugar from sugarcane or from sugar beets. For instance, if the FS product replaces sugar from sugarcane, the effects on the human toxicity and ozone related categories decrease to the point that the impact is considered negative, i.e. there are benefits related to these categories. Decreases in impact categories when FS product replaces sugar beets are more modest, however categories such as ozone depletion and formation, and human toxicity do show improvements. Furthermore, in combination with Table 54, it can be seen that the impact on the categories in which conventional palm oil production has highest repercussions (i.e. aquatic eutrophication, human toxicity and terrestrial eutrophication) is greatly reduced. Looking at the overall single score, by producing MICCI oil with FS products replacing sugar from sugarcane or sugar from sugar beet, the palm oil impact assessment score is decreased from 4.6 weighted score points, to -1.6 and 2.8 total scores respectively, see Figure 20.

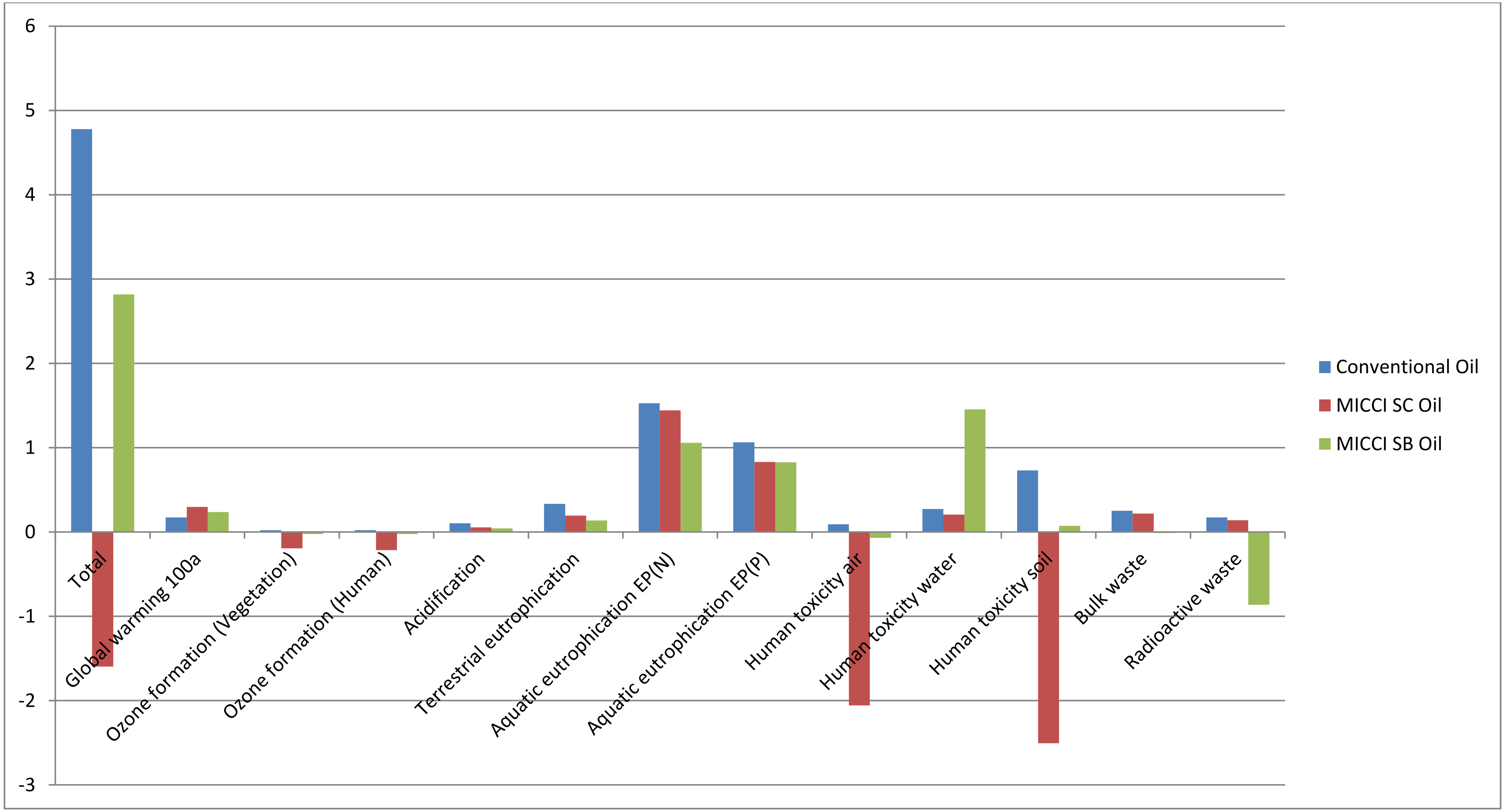

Figure 20. Comparison of conventional palm oil vs MICCI palm oil, Impact Assessment Results of Palm Oil
Values represent EDIP2003 single score points. SC – FS Product as sugar from sugarcane replacement; SB - FS Product as sugar from sugar beet replacement.

10. Conclusions

In this study the feasibility of 2nd generations intermediates obtained from oil palm residues has been evaluated. After a comparative study with bio-oil, bio-ethanol and lactic acid, it was found that a fermentable sugars product is the most promising alternative for a 2nd generation palm-based biorefinery. This selection was based on the added value reached by transforming the biomass residues into sugars, and the logistic, safety and environmental impact it adds to a palm oil mill.

As part of the process design for the biomass transformation, it is clear that a pre-treatment step is necessary due to the recalcitrant nature of the material. It is found that acid catalysed steam explosion is the most suitable alternative, based on several criteria, such as high conversions and low product degradation. Also, the suitability of the pre-treatment alternative for integration into the palm oil mill setting is considered, finding that other alternatives require more complex processes for detoxification or catalyst and/or lignin recovery. The resulting cellulosic fraction undergoes enzymatic hydrolysis to give the fermentable sugars. Additionally, when biomass available form plantation operations is processed, additional sugars are recovered from milling these raw materials.

A preliminary evaluation of the process indicates that the benefits related to economies of scale are overwhelmed by the biomass transportation costs that larger scales imply. Under the given assumptions for the preliminary evaluation, an optimum scale of four mills (biomass derived from operations of four palm oil mills with capacity of 60 tonne/h each) was found. Moreover, when biomass from the plantations related to the mills is also processed, the optimum scale is three mills. This difference is expected, as the processing of biomass from plantations to a central mill implies more transport expenses.

Considering the proposed process, 357 and 1 194 ktonne/y of fermentable sugars are produced from processing biomass residues from three and ten mills respectively. This product stream is the result of lignocellulosic sugars conversion, as well as OPT and OPF milling. Energy demands of the process, including biomass pretreatment and fermentable sugars product heat treatment, are covered by combined power and heat generation from unconverted solid residues. The total generated MP steam is consumed in the process, while exceeding LP steam is considered as a by-product covering the steam requirements of the adjacent mill.

The proposed process requires a total capital investment of 536 and 1 150 million RM when processing biomass derived from three and ten mills cases respectively. The main capital investment contributions are related to the pretreatment and the CPH systems. Once the capital investment and other direct and indirect cost are taken into account, the proposed design yields a fermentable sugars product with total processing costs of 1 074 and 1 120 RM/tonne of sugar for the three- and ten-mill cases respectively, reflecting the large impact of transportation costs. As larger scales imply higher selling volumes, the project at ten-mill scale case has higher NPV. However, when the amount of fermentable sugars produced in the project life is considered, the present value of the project is slightly higher for the three-mill case, with 368 RM/tonne of sugar produced *vs.* 356 RM/tonne sugar for the ten-mill case.

Additionally, the three-mill case is clearly more robust to changes on transportation distance amongst mills. In the ten-mill case, production cost reach the LTC sugar price when the distance is increased *ca.* 1.8 times the average distance when the mills and plantations are adjacent to each other. In the three-mill case, this situation is reached when the distance is increased more than 4 times. On the other hand, changes in sugar prices greatly affect the project at any scale, decreasing the specific NPV around 35% and 70% for both cases when the sugar price is 15 and 30% lower than the LTC sugar price respectively.

Moreover, the environmental benefits related to the integration of the process in the oil palm production are identified through an LCA following the EDIP 2003 method. It is found that categories in which conventional palm oil production has the largest repercussions, i.e. aquatic and terrestrial eutrophication, and human toxicity, are reduced by the integration of the MICCI process with fermentable sugars as by-product.

Thus, the proposed design is an initial set-up delivering a profitable fermentable sugars product that brings benefits to the palm oil production chain. Considering that the capital requirements are higher for the ten-mill case, and that the profitability of the project is more sensible to changes in transportation distance and sugar prices at higher scale, it is considered that the project at a three-mill case is a better alternative.

11. Recommendations

Given the scope of the present project, there are several points that were taken into account based on literature data or estimations. In order to better tune this study, it is recommended that following works put special attention to the next points:

- Biomass composition: In the present study the composition of biomass is taken from literature data, which in some cases is incomplete or different amongst authors. In order to verify the biomass composition and changes related to plantation age, it is recommended to analyse the biomass residues from different ages and identify components that could vary significantly. It is also recommended that composition analysis differentiate the main components of hemicellulose and lignin to have a more accurate estimation of their conversion or degradation.

- Pretreatment and enzymatic hydrolysis: Conversion yields used in the present study are based on experimental data with other agricultural residues. Although attainable yields are similar for most agricultural residues, oil palm biomass residues differ on their composition and microstructure, and thus it is considered relevant to verify the effect of the pretreatment on the structure of these residues.

- Process parameters: It is considered that parameters related to sugar recovery should be optimized. A straight forward parameter for optimization is water used for imbibition, which can be estimated by comparing the cost of transporting a more dilute product or product concentration *vs.* attainable sugar recovery. Other parameters, such as temperature and retention time for pretreatment and enzymatic hydrolysis are considered important and can be improved with kinetics data related to oil palm biomass.

- Transportation of biomass: In the present design project, an average distance for transportation is assumed. Considering the impact this factor has on the profitability of the project, it is considered relevant to verify whether considered mills and plantations are adjacent to each other.

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APPENDICES

APPENDIX 1. Product alternatives report

1. Product alternatives

This report is part of the exploratory design project that investigates the techno-economic feasibility of 2nd generation intermediates production from solid biomass (BM) available at palm oil mills in Malaysia.

At the mills, fresh fruit bunches (FFB) from oil palms are processed for the extraction of oil and consequently, empty fruit bunches (EFB) are obtained as a by-product. Shells, fibres, and kernels are also produced after the fruit processing, although in smaller scale when compared to EFB (table 1). Additionally, other biomass by-products such as oil palm fronds (OPF) and oil palm trunks (OPT), are constantly generated at the plantation sites (Agensi Inovasi Malaysia, 2011). In the 2nd generation palm-based biorefinery, compatible to the framework of a conventional palm oil mill structure, it is possible to transform these solid biomass residues into higher value products. In this preliminary report, a final product of the envisioned 2nd generation palm-based biorefinery will be selected.

Biorefinery is a process concept that focuses on resource optimisation to produce a variety of fuels, chemicals, feed, materials, and energy from biomass. As such, a biorefinery has as immediate target the production of chemical building blocks or intermediates that can be further converted into a larger number of products inside their own structure, or exported to a partner company that specializes in the intermediate further conversion. In order to select the target product, or the point until which the biorefinery will be involved, a market assessment must be made considering the value chain related to the product and identification of business partners. Thus, the palm-oil mill as a biorefinery has the possibility to extend its scope to produce besides energy, fuels, bio-based chemical intermediates, or higher-value bio-based chemicals from available biomass.

For this project, four major alternatives for solid biomass conversion and integration in the palm mill product portfolio are studied, namely bio-oil, fermentable sugars, bio-ethanol and lactic acid. In order to select the most appropriate alternative, the four products and their production implications are going to be compared based on the six following criteria:

- High economic benefits
- Low complexity
- Diverse market opportunities
- Sufficient market size

- Environmental implications
- Storage and handling implications

Low complexity level of the process is required so that the design can be implemented in already existing facilities. Additionally, simplicity is very relevant considering the remote location of the mills, where extra services might not be readily available. The economic benefit is to be based on a potential economic margin, obtained by subtracting the cost of the main inputs, mainly biomass cost, to the possible revenues from each alternative. Additionally, the economic margin potential is converted to money flows, based on a 60 t/h palm oil mill with an annual operation time of 6000 h.

Calculations are based on biomass available from FFB processing in a conventional mill as reported by the (Malaysian Department of Environment, 1999) and (Yusoff, 2006). Composition of EFB, shell and fibres is taken as reported by (Shamsudin *et al.*, 2012) and (Kumar *et al.*, 2010), see table 1. Only mill-derived biomass is considered for this initial comparison, since the inclusion of OPF and OPT would not significantly change the economic potential comparison of the product alternatives.

Table A1-1. Composition of palm oil biomass residues

Component	EFB	Shell	Fibres
Available biomass (%FFB)	23	6	13.5
Cellulose (dry weight %)	36	25	36
Hemicellulose (dry weight %)	25	21	24

Biomass costs are calculated according to the National Biomass Strategy published by the (Agensi Inovasi Malaysia, 2011).

1.1. Bio-oil

Pyrolysis is the process of degradation when there is no or limited oxidizing agent (e.g. oxygen in air), and usually occurs at temperatures in the range of 500 – 800 °C. Pyrolysis of biomass yields char, gas, and liquid bio-oil, which has a much higher density than woody materials. Bio-oil is attractive for co-firing, or it can be converted to transportation fuel through upgrading processes (FAO, 1994). Other products, such as furans and levoglucosan, can be recovered from bio-oil depending on the pyrolysis conditions. Different approaches have been designed to favour the production of a desired pyrolysis product. In the case of bio-oil, fast and flash pyrolysis have been employed to favour its production.

Biomass pyrolysis research has been mainly focused on woody biomass with relatively low moisture, such as timber wood. Prior to the pyrolysis process, the biomass is further dried to reach moisture contents of around 7% (Ringer *et al.*, 2006). Commercialisation plans for biomass pyrolysis with bio-oil as product, take the gas and char produced to supply part of the energy required for drying the incoming biomass (Cole Hill Associates, 2004).

Bio-oil can be used for heat and power purposes, mainly in applications such as substitute for boiler fuel, district heating heavy fuel and gas, as well as fuel for kilns. Furthermore, pyrolysis oil can replace diesel in stationary industrial engines, although it cannot be used in transportation without upgrading. Therefore, an indication of its price in the market is based on its heating value when compared to that of heating oil and its price. Bio-oil also holds the possibility to be upgraded to transportation fuels or converted to other chemicals, similar to crude oil. To have an upper value for bio-oil, the crude oil price and its energy content are used a higher reference.

1.1.1. *Assumptions*

- Pyrolysis yields are based on numbers given by (Cole Hill Associates, 2004) for fast biomass pyrolysis based on dry basis: 70% liquid, 17% char and 13% gas products.
- Bio-oil is considered as the main valuable product. Its price is based on a heating value of 18 MJ/L, being 47% of heating oil (Ringer *et al.*, 2006).
- Considering the Malaysian price for heating oil as 2.51 RM/L (Indexmundi, 2012), the lower value (LV) of bio-oil is taken as 1.16 RM/L.
- The high value bio-oil (HV) is based on the highest crude oil price in the period 2009 - 2012, at 2.26 RM/L[1], (Indexmundi, 2012).

1.1.2. *Results*

Table A1-2. Bio-oil estimations

Bio-oil production (t/t dry BM)	0.70	
Bio-oil cost (RM/t dry BM)	163	
Biomass cost (RM/t dry BM)	163	
Market price	LV	HV
Bio-oil revenue potential (RM/t dry BM)	658	860

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[1] Taking the crude oil price reported for March 2012, 358.59 RM/bbl.

Economic margin potential (RM/t dry BM)	495	697
Economic margin potential ($1x10^6$ RM/y)	37	52

1.2. Fermentable sugars

The availability of renewable feedstock has been identified as a key aspect to incentivize the bio-based chemical industry growth in Malaysia. Moreover, fermentable sugars have been recognised as the largest feedstock that can supply the growing bio-based chemical industry (National Research Council, 2000). This is supported by the fact that sugars are the initial component of plenty of well-established fermentations for production of commodities and fine chemicals (e.g. citric acid, ascorbic acid, etc.) and holds a large potential in emerging bio-based products. In this context, the production of monomeric sugars derived from already available palm oil solid waste is an appealing alternative.

Fermentable sugars, as an intermediate product, are not commonly found on the market as such. Molasses, a sugar by-product sometimes used for industrial fermentations, and unrefined raw sugar can be taken as an indication of the fermentable sugars market. In Malaysia, the sugar industry imports most of its raw material; the largest sugar refiner imports 97% of its raw sugar from Brazil and Australia (CIMB Reseach, 2011). Furthermore, sugar prices have been historically regulated in Malaysia, and in early 2012 a new long-term contract (LTC) was signed for raw sugar supply into the country. The new contract fixes the raw sugar supply price at 1.77 RM/kg until 2014, with the objective to mitigate the volatility of sugar markets (Ariffin, 2012).

Considering the market volatility and the relative short duration of the LTC, in this report sugar price is taken in a range from the lowest and highest raw sugar prices in the international markets in the past 3 years. Inside this range is the LTC price mentioned before, 1.77 RM/kg, as well as reported molasses price (based on sugar content) in Malaysia, 1.11 RM/kg (CIMB Reseach, 2011).

1.2.1. *Assumptions*

- The bio-chemical conversion into fermentable sugars considers the following reactions:

$(C_6H_{10}O_5)_n + n\ H_2O \longrightarrow n\ C_6H_{12}O_6$ Cellulose hydrolysis

$(C_5H_8O_4)_n + n\ H_2O \longrightarrow n\ C_5H_{10}O_5$ Hemicellulose hydrolysis

- The maximum theoretical conversion yields are based on the previous equations, with 1.11 g glucose per g cellulose and 1.14 g xylose per g hemicellulose. Both pentoses and hexoses are considered fermentable sugars.
- Pre-treatment cost contribution is based on acid hydrolysis treatment, with the NREL report by (Humbird *et al.*, 2011) as reference. The acid cost impact is based on 0.022 tonne H_2SO_4 per tonne of dry biomass.
- The enzyme cost impact is the same as described in section 1.3.1 for ethanol, considering that the same amount of polymeric sugars is hydrolysed enzymatically for both product alternatives.
- Sulphuric acid price is 0.27 RM/kg (ICIS Pricing, 2012)[2].
- Fermentable sugars price is evaluated considering 3 values: The lowest and highest raw sugar prices in the period May 2009 – May 2012, 1.05 RM/kg and 2.32 RM/kg respectively[3], and the Malaysian LTC raw sugar supply price, 1.77 RM/kg.

1.2.2. *Results*

Table A1-3. Fermentable sugars estimations

Sugars production (t/t dry BM)	0.64		
Biomass cost (RM/t dry BM)	163		
Pretreatment cost (RM/t dry BM)	6		
Enzyme cost (RM/t dry BM)	126		
Sugar market price	LTC	LV	HV
Sugar revenue potential (RM/t dry BM)	1136	677	1490
Economic margin potential (RM/t dry BM)	841	383	1195
Economic margin potential (1x10^{6} RM/y)	63	29	90

1.3. Bioethanol

The process from polymeric carbohydrates, the main components of the available biomass, to ethanol requires hydrolysis of biomass into pentoses (xylose) and hexoses (glucose), similar to the fermentable sugar case. Furthermore, the hydrolysis product needs to undergo a fermentation step to produce ethanol. The resulting ethanol is obtained in a diluted medium and thus requires extensive downstream processing to reach set market demands (e.g. concentration by distillation). There is ample literature regarding bio-ethanol production and several integrated fermentation schemes have

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[2] Conversion from Malaysian Ringgit to US dollars (**0.324 USD/RM**) and to euros (0.25 EUR/RM) is based on the average conversion rates in the period from November 2011 to May 2012. This conversion rate is used throughout the whole report.

[3] Prices based on ICE trade values for contract no. 11 sugar, with lowest price taken from May 3rd, 2010 and the highest price taken from January 31st, 2011, taken from (Barchart 2012).

been developed, such as simultaneous saccharification and fermentation. Even more integrated processes have been developed, such as the consolidated bioprocessing with saccharification and fermentation of C_5 and C_6 sugars taking place at the same time as saccharolytic enzymes are produced.

Ethanol as fuel accounts for 75% of global ethanol use. Furthermore, fuel ethanol is under current growth in Asia, where blending applications are dominant. On the other hand, downstream applications in Asia are mainly related to acetic acid, ethyl acetate and as solvent in pharmaceutical products. Additionally, hydrous industrial ethanol or Brazilian B-grade ethanol is currently traded in Asia (ICIS Pricing, 2011).

1.3.1. *Assumptions*

- Sugar yield is based on the reaction presented in section 1.2.1.
- The monomeric sugar fermentation into ethanol follows the chemical reaction:

$$C_6H_{12}O_6 \xrightarrow{\quad} 2\ CH_3CH_2OH + 2\ CO_2$$ Glucose fermentation

$$C_5H_{10}O_5 \xrightarrow{\quad} 5/3\ CH_3CH_2OH + 5/3\ CO_2$$ Xylose fermentation

- Calculations are based on the maximum theoretical yield (based on the reaction equations): 0.511 gram of ethanol per gram of xylose or glucose.
- Ethanol price is also taken as a range with 2.84 RM/kg as lower value, based on Asian market prices for fuel grade ethanol (Argus media Ltd., 2011), and 3.32 RM/kg as higher value based reported B-grade ethanol prices (ICIS Pricing, 2011).
- Enzymes cost is based on the NREL report for lignocellulosic ethanol (Humbird *et al.*, 2011), with enzymes accounting for 0.382 RM/L of ethanol produced[4].

1.3.2. *Results*

Table A1-4. Bioethanol estimations

Bio-ethanol production (t/t dry BM)	0.33	
Biomass cost (RM/t dry BM)	163	
Pretreatment cost (RM/t dry BM)	6	
Enzyme cost (RM/t dry BM)	126	

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[4] Converting 2007 dollars to 2011 dollars using CPI values reported at the US Bureau of Labour Statistics: ftp://ftp.bls.gov/pub/special.requests/cpi/cpiai.txt

Market price	LV	HV
Ethanol revenue potential (RM/t dry BM)	932	1090
Economic margin potential (RM/t dry BM)	638	795
Economic margin potential (1x10^6 dry RM/y)	48	60

1.4. Lactic acid

Lactic acid is an organic acid commercially produced by fermentation. It is used as an ingredient in the foods and cosmetics industry, as well as additive in textiles, films and foams manufacture. Recently it is best known as an attractive compound that can be used for the production of polylactic acid, a replacement for polyethylene terephthalate (PET) in the plastics sector.

Lactic acid is typically produced through sugar fermentation, with a characteristic high yield from glucose. However separation and purification steps are still regarded as expensive, especially considering the diluted form in which it is produced during fermentation, similar to bioethanol. Additionally, it is common to generate one tonne of gypsum per tonne of lactic acid as waste. Several purification methods have been used in the industry and membrane based technologies seem to be giving the most cost efficient results, however membrane replacement still have a significant contribution to operating costs (Janssen, 2012).

1.4.1. *Assumptions*

- Pretreatment and hydrolysis yields are the same as those used in the fermentable sugars and ethanol sections of this report.
- The monomeric sugar fermentation into lactic acid follows the chemical reaction:

$C_6H_{12}O_6 \longrightarrow 2\ CH_3CHOHCOOH$ Glucose fermentation

$C_5H_{10}O_5 \longrightarrow CH_3CHOHCOOH + CH_3COOH$ Xylose fermentation

- Calculations are based on the maximum theoretical yield (based on the reaction equations): 1 gram per gram of glucose and 0.6 g per gram of xylose.
- Pretreatment and hydrolysis costs are considered the same as for bioethanol.
- Lactic acid price is presented in a range based on market price indications (ICIS Pricing, 2012), taking 6.57 RM/kg for technical grade as lower value, and 8.50 RM/kg for food grade as higher value.

1.4.2. *Results*

Table A1-5. Lactic acid estimations

Lactic acid production (t/t dry BM)	0.54	
Biomass cost (RM/t dry BM)	163	
Pretreatment cost (RM/t dry BM)	6	
Enzyme cost (RM/t dry BM)	126	
	LV	HV
Lactic acid revenue potential (RM/t dry BM)	2897	3518
Economic margin potential (RM/t dry BM)	2603	3223
Economic margin potential ($1x10^6$ dry RM/y)	195	242

1.5. Added value to biomass

The price of the products of a biorefinery has an inverse relation to the volume or amount produced, which is a result of the processing steps required to obtain them. The previous statement is reflected in figure 1. The lower blue marks of the graph represent the accounted costs, namely for biomass and pre-treatment. The upper marks of the bars, in green, represent the different market prices of the respective product, thus the economic margin is the difference between the cost (blue marks) and the price (green marks). As mentioned in the previous section, price ranges are considered depending on possible markets niches for the product. The economic margin per tonne of product is higher for the more processed and refined products, with bio-oil in the lower value extreme and lactic acid in higher extreme.

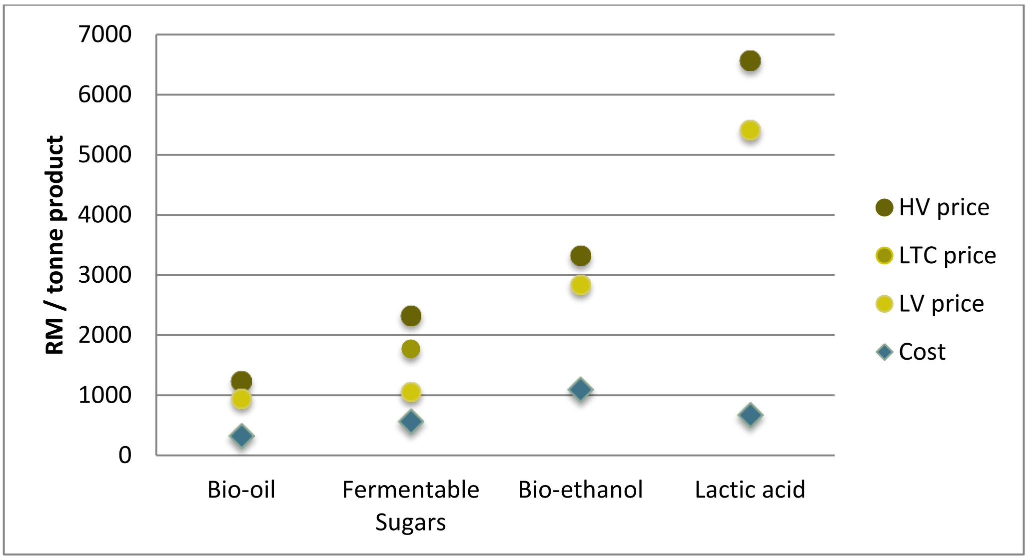


Figure A1-1. Value of the four product alternatives
HV: high value, LV: low value, LTC: Long term contract.

However, as conversion yields are far from 100% and different amount of products can be obtained from the same biomass, the economic value per tonne of product is not an appropriate point of comparison. The economic margin is best accounted on a biomass basis, which can be seen as the added value to the biomass, see figure 2. It is clear that lactic acid production brings the most value to biomass, being in a higher position in the value chain of bio- based chemicals. However, the presented values are based on potential production without regard on purification costs, which in the case of both ethanol and lactic acid have been reported to have a major contribution to the total processing costs.

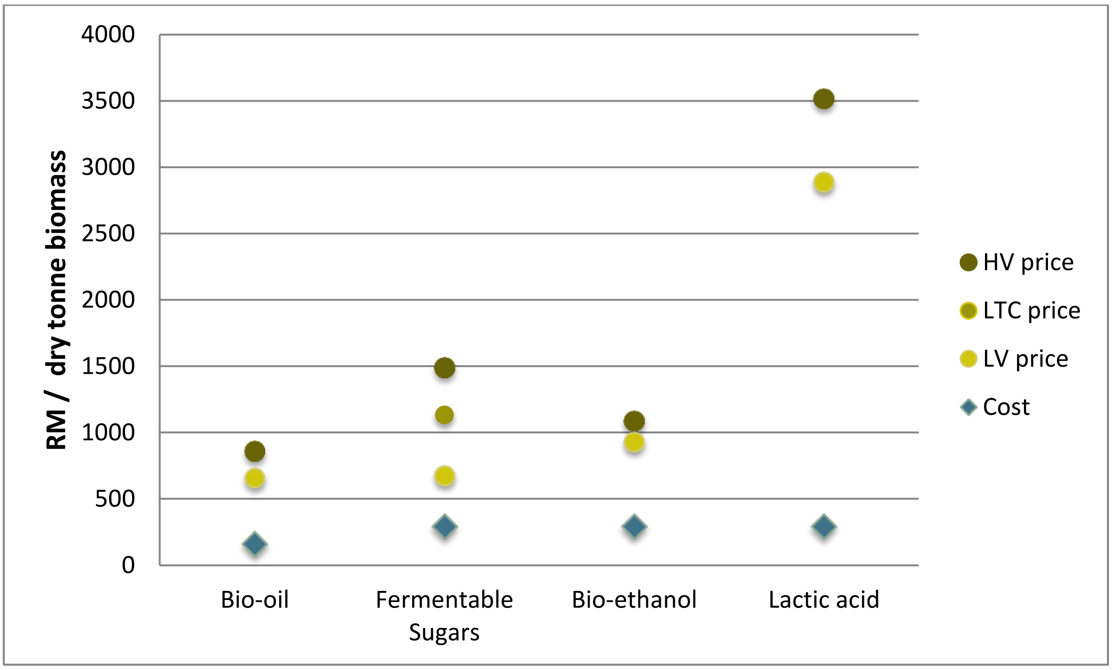


Figure A1-2. Value of the product alternatives per tonne of processed biomass
HV: high value, LV: low value, LTC: Long term contract.

The range of possible economic margin or added value achieved with production of fermentable sugars is quite ample. Bearing in mind the molasses price found in the literature, which is similar to LTC price, the economic margin is comparable to that of bioethanol and higher than bio-oil. The mentioned molasses price however, corresponds to 2010 when raw sugar prices were low. In view of the higher current price of raw sugar compared to 2010, it is expected that molasses price in the Malaysian market is higher in 2012 and will stay in the same higher level at least for the duration of the LTC deal.

After considering the market situation for raw sugar, the results obtained with fermentable sugars can be taken as more favourable than those with bio-ethanol. Bio-oil on the other hand, brings little value to the biomass, while lactic acid yields the most added value.

2. Discussion and selection

In order to select the product alternative based on the palm mill scenario, the four alternatives are compared to each other with the specified criteria as basis. The score of the product alternatives for each criterion is summarized in Table 6.

Table A1-6. Comparison of different product alternatives based on selection criteria.

	Bio-oil	Fermentable sugars	Bio-ethanol	Lactic acid
High economic benefits	-	+	0	+
Diverse market opportunities	-	+	-	+
Market size	+	0	+	0
Low process complexity	0	+	-	-
Storage and handling implications	-	0	0	0
Environmental implications	-	+	0	0

Positive (+), negative (-) or neutral (-) compared to each other.

Based on the estimated economic margin potential presented above, lactic acid integration in the oil palm mill product portfolio seems the most attractive option (*ca.* 3000 RM/t dry biomass compared to 400-1200 RM/t biomass for sugars, 650-800 RM/t biomass for bio-ethanol and 500 – 700 RM/t biomass for bio-oil). Regarding market opportunities, lactic acid has entry points to different value chains, ranging from food and pharmacological products to bio-plastics. On the other hand, fermentable sugars as a product could be placed as intermediate for the chemical industry, since they are the starting point of numerous fermentation processes, and thus help to cover the local supply for the growing chemical industry in Malaysia. Furthermore, fermentable sugars have a potential use for the production of a wide range of emerging bio-based commodities and fine chemicals. In this context, the palm oil mill would remain in the role of intermediates producers, delivering CPO, kernels and fermentable sugars ready for further processing.

Bio-ethanol is well established in the growing bio-fuels market. Entering the extensive fuels market would have to be a new initiative from the palm oil mill sector, although it is being achieved in certain cases, and to certain degree, with biodiesel from palm oil. However, the ethanol biofuel market is strongly dominated by Brazil, the US, and more recently China, where ethanol selling price is supported by decades of technology and infrastructure development, large scales and/or government schemes and subsidies. Moreover, in these countries ethanol is produced from starchy biomass which imply less economic burdens as no extensive pre-treatment is necessary (Walter *et al.*, 2007).

Bio-oil also fits in the biofuels market, however its use cannot be set as transportation fuel unless it undergoes upgrading processes. Overall, bio-oil is restricted to the heating-oil market niche, for use in boilers and kilns mainly, carrying a lower heating value than standard heating and crude oils. As mentioned before, bio-oil can be upgraded, however efficient technologies are still being developed to increase yields and facilitate the process. It has been reported that around 40% of pyrolytic lignin (*ca.* 30% of bio-oil) can be recovered as transportation fuel (Marker *et al.*, 2005). However, the operations necessary for this are more related to the petroleum infrastructure, like hydrocracking and reforming, where necessary hydrogen is available.

Process complexity related to fermentable sugars, ethanol and lactic hold certain similarities. Production of fermentable sugars, ethanol and lactic acid products require pre-treatment and hydrolysis steps. However, fermentable sugars production might only require an additional concentration step to achieve a desired specification, whereas ethanol and lactic acid production requires more extensive fermentation and downstream processing. Although hydrolysis and fermentation could be integrated into a single step, its implementation might require finer process control, and the tighter safety standards that any modified microorganism implies. On the other hand, pyrolysis process for biomass holds the disadvantage of requiring the removal of large amounts of water. In the case of EFB, with a typical moisture content higher than 50%, the energy use for water removal can amount to a significant economic and ecological burden.

As for safety and handling risks, both ethanol and bio-oil have in common flammability as a safety risk. Even more, bio-oil is highly corrosive to most metals and thus must be transported and stored in stainless steel containers. Bio-oil, compared to other fuels, is not very stable as undesired reactions amongst the compounds that conform it still take place after processing, especially at higher temperatures. As a result, heavy weight tars, sludge, waxes and water phases are formed and viscosity increases in time. The practical result of these property changes is the relative short shelf-life bio-oil has, lasting around four weeks (Ringer *et al.*, 2006). This is a significant characteristic of bio-oil, which might affect the mill and its logistics. Bio-ethanol, as a volatile compound, also implies caution in transportation and storage, although not as strict as bio-oil. In contrast, lactic acid is less reactive and flammable, although highly acid and irritant, ranked with higher health risks than ethanol. Fermentable sugars, on the other hand, are quite mild in comparison with no regarded health nor environmental hazard. If stored at high concentrations, sugars are very stable for prolonged periods of time. However, at lower concentrations, fermentable sugars are subject to microbial activity and might thus require proper preventive measures to avoid deterioration.

All the four alternatives are fossil-fuel substitutes and as such will be regarded to hold environmental benefits when considering greenhouse gases emissions. However, in plant mill operations, bio-oil of the four alternatives is considered to hold a higher environmental risk as an acidic and corrosive liquid, although it is more biodegradable and represents a smaller ecological hazard than petroleum oils. On the other hand, bio-oil's complex composition has given rise to concerns on human health effects, specially carcinogenicity derived from benzene and phenanthrene (Ringer *et al.*, 2006).

Altogether, fermentable sugars are the most promising product alternative for integration to the product portfolio of existing palm oil mills. Hence, in the further work of this exploratory design project, fermentable sugars will be taken as the final product of the 2nd generation palm-based Biorefinery in Malaysia.

3. Additional information. Mill and plantation biomass

As explained before, for each processed FFB not only biomass residues are generated at the mill, but also OPF and OPT. As these materials are commonly left on the plantation site for nutrient recycle, their use would clearly imply increased biomass cost due to their transportation.

Biomass costs are calculated according to the National Biomass Strategy published by the (Agensi Inovasi Malaysia, 2011). The price allocation must consider four different cost: (1) substitution, (2) harvesting plus collection, (3) pre-processing, and (4) transportation. Substitution cost is considered as replacement cost for EFB, OPT, and OPF, whereas for shells and fibres it is based on market price. Harvesting and transportation costs are not considered for EFB, fibres, and shells as they are readily available at the mill. Pre-processing cost, mainly attributed to drying as mentioned in the National Biomass Strategy, is not taken in the calculations since if required, it is inside the battery limits. The list of costs is presented in table 7, taking the average of values given in the mentioned strategy report. Transportation cost of OPF and OPT is based on a plantation area of 18 000 ha and considering an average distance. Assuming a circular plantation area with a mill at the centre, the average distance is taken as 2/3 of the radius, and is thus 5.05 km.

Table A1-7. Biomass costs for biomass, in RM/tonne wet-basis.

Biomass type	Substitution	Harvesting	Transport	Total
EFB	93	0	0	93
OPF	35	12	26	73

OPT	78	14	26	118
Shell	130	0	0	130
Fibre	40	0	0	40

3.1. Fermentable sugars

3.1.1. *Assumptions*

- Mill biomass is taken as in section **Error! Reference source not found.** of the main report.
- The plantation area related to the mill is taken as 18 000 ha, considering FFB yields of 19.7 tonne/yr/ha as reported by the (Malaysian Palm Oil Board, 2012).
- OPT and OPF are available at 0.15 and 0.55 dry tonne/tonne FFB respectively, based on replanting and pruning rates, given by (Yusoff, 2006).
- Only the basal part (1/3) of OPF is considered, based on the 2012 NBS (Agensi Inovasi Malaysia, 2011).
- Composition and moisture content of OPF and OPT is taken as 70%, based on published data by (Zahari *et al.*, 2012) and (Kosugi *et al.*, 2010), see table 8.
- Biomass costs are based on numbers published in the 2012 NBS, taking transportation cost only for plantation derived biomass.
- Transportation distance is based on the average transportation distance, assuming a circular plantation area with a mill at the centre.

Table A1-8. Composition of OPT fibres and sap

Component	OPT	OPF
Fibre	**g/kg DW**	
Cellulose	412	417
Hemicellulose	344	164
Lignin	171	155
Ash	73	264
Sap	**g/L**	
Fructose	3	2
Glucose	52	54
Sucrose	3	21

- Fibre conversion into sugars is described in section 1.2.1 of the main report.

3.1.2. *Results*

Table A1-9. Fermentable sugars estimations with biomass from mill and plantation

Sugars production (t/t dry BM)	0.80		
Biomass cost (RM/t dry BM)	254		
Pretreatment cost (RM/t dry BM)	6		
Enzyme cost (RM/t dry BM)	157		
Sugar market price	LTC	LV	HV
Sugar revenue potential (RM/t dry BM)	1418	845	1859
Economic margin potential (RM/t dry BM)	1000	428	1442
Economic margin potential (1x10^6 RM/y)	195	84	281

3.2. Multiple mills

3.2.1. *Assumptions*

- Two scales are evaluated, considering biomass derived from 2 and 5 mills. Transportation costs implications are based on plant locations as presented in figure 1.
- For multiple mills, the plantation area is considered 2x and 5x times of a single mill area. Transportation of OPF and OPT is based on the average distance as explained before. Additionally, it is considered that 1/2 and 4/5 of the mill derived biomass (EFB, shells and fibres) from 2 and 5 mills respectively, are to be transported to a central mill. It is taken that distance amongst mills is the average distance from any point in the total plantation area to the centre.

Table A1-8. Biomass costs for biomass, in RM/tonne wet-basis.

Biomass type	Substitution	Harvesting	Transport	Total
EFB	93	0	18	111
OPF	35	12	36	84
OPT	78	14	36	128
Shell	130	0	18	148
Fibre	40	0	18	58

Table A1-9. Biomass costs for biomass, in RM/tonne wet-basis.

Biomass type	Substitution	Harvesting	Transport	Total
EFB	93	0	46	139
OPF	35	12	58	105
OPT	78	14	58	150
Shell	130	0	46	176
Fibre	40	0	46	86

3.2.2. *Results*

Table A1-10. Fermentable sugars estimations from multiple mills and plantations.

	Mill			M+P		
	LTC	LV	HV	LTC	LV	HV
	Single mill					
Economic margin potential (RM/t dry BM)	841	383	1195	1001	428	1442
Economic margin potential ($1x10^6$ RM/y)	63	29	90	195	84	281
	Two mills					
Economic margin potential (RM/t dry BM)	805	346	1158	965	392	1406
Economic margin potential ($1x10^6$ RM/y)	121	52	174	373	153	548
	Five mills					
Economic margin potential (RM/t dry BM)	749	290	1102	900	327	1303
Economic margin potential ($1x10^6$ RM/y)	281	109	413	877	319	1308

M+P – Mill and plantation related biomass

4. Net present value

4.1.1. *Assumptions*

- The lower price from the ranges used in the previous sections is used for calculating the NPV using the following equation:

$$NPV := \sum_{(n)} \frac{nrev_s \cdot (1 - tax)}{(1 + i)^n}$$

where:

$nrev_s$ = economic margin (1×10^6 RM/y),

tax = corporate tax in Malaysia, taken as 0.25 (Malaysian Investment Development Authority, 2012)

i = interest rate in Malaysia taken as 0.065, based on the highest lending rate reported in 2011 and that in the first quarter of 2012 (Central Bank of Malaysia, 2012).

4.1.2. *Results*

Table A1-11. NPV results for different scales, in millions of RM.

Biomass	1 mill	2 mills	5 mills
Mill	160	280	590
M+P	450	820	1720

M+P – Mill and plantation related biomass

APPENDIX 2. Pretreatment alternatives

Different pre-treatment methods and their combinations have been developed, each with different results on each of the key factors mentioned above. In this design project, importance is given to the production of monomeric sugars from both pentoses and hexoses with high yields. A simple process that can be easily adapted into the palm oil mill setting is preferred.

Below a brief description of the methods most available in the literature is presented. The descriptions are based on publications by (Takara *et al.*, 2010), (Alvira *et al.*, 2010), (Kumar *et al.*, 2009) and (Mosier *et al.*, 2005).

1. Description of pre-treatment technologies

1.1. Biological

This comprises the conversion of the lignocellulosic material structure with the use of microorganisms like brown, white, and soft-rot-fungi that degrade lignin and hemicellulose. Although highly specific and with little or no formation of inhibitors, biological pre-treatment has the major drawback that the hydrolysis rate is very low when compared to other technologies (Mosier *et al.*, 2005).

1.2. Mechanical comminution

Mechanical comminution is a pre-treatment method that reduces the size of the biomass size. Besides a degree of crystallinity reduction is achieved, depending on the final particle size. Common particle sizes are 10-30 mm after chipping and 0.2-2 mm after milling or grinding. Power consumption and resulting costs, may become high depending on the final particle size (Alvira *et al.*, 2010). This method commonly used before other pre-treatment methods in order to achieve overall high yields.

1.3. Steam explosion

For this treatment, high pressure steam is put in contact with biomass and as the pressure is lowered rapidly, the structure of the lignocellulose material is greatly affected. This pretreatment has two main effects: a mechanical opening of the fibres due to the rapid decompression, and acetic acid formation from released acetyl groups present in hemicellulose that favour its further hydrolysis (Mosier *et al.*, 2005). To improve its effect on hemicellulose hydrolysis, external acid addition has been considered, however this might result in larger capital expenditure related to equipment requirements (Alvira *et al.*, 2010)).

(Shamsudin *et al.*, 2012) published a study on steam explosion of EFB obtained after common mill threshing operations, obtaining 209.4 g of reducing sugars / kg EFB. This treatment seems rather promising as no catalyst was needed, only steam already available at the mill.

1.4. Acid treatment

In this case, the hemicellulose fraction is solubilized by the presence of an acid agent, which results in release of oligomers that are further converted to monosaccharides and can further be converted to furfural and HMF. Both diluted and concentrated acid treatments have been used, the first one more commonly preferred due to lower equipment requirements and lower formation of inhibitors which is largely dependent of the severity of the treatment (Alvira *et al.*, 2010). Dilute acid pretreatment, usually at concentration below 4%, successfully hydrolyses hemicellulose to xylose (xyl) at high reaction rates, making cellulose more available for hydrolysis to glucose (glc).

This treatment method as a general rule can be applied under two different conditions, at high temperatures with low solids loadings (5 – 10%) or at low temperatures for higher solids (10-40%). It has been proven successful for treatment of different plant materials like hardwood, grasses and stover, among others.

The diluted acid pre-treatment of corn stover, with cellulose, hemicellulose and lignin content similar to EFB (Mosier *et al.*, 2005), has been widely studied and reported in the literature. Optimum conditions for this material have been found to be 2.0% sulphuric acid with reaction time of 43 min at 120 °C, resulting in 77% xylose conversion and 8.4% of glucose yield (Kumar *et al.*, 2009).

Depending on the severity of the treatment, degradation of C_5 and C_6 sugars might result in lower yields as well as formation of toxic compounds that are inhibitory to microorganisms used in fermentations. To limit the formation of inhibitors without losing sugar yields, dilute hydrolysis can be performed at mild conditions in a two-step approach. In the first stage, a very short residence time (5-10 min) allows the conversion to only release xylose oligomers. Then, in a second step with lower temperature and longer residence times the oligomers are further converted into monomers without significant degradation (Humbird *et al.*, 2011).

Using the two-stage approach (with oligomer conversion) resulted in xylan to xylose yields up to 80% (90% cellulose to glucose yield after enzymatic hydrolysis). However, little improvement was achieved with this process and further research has been focused on using hemicellulases together with cellulose hydrolysis, avoiding the need of a second step for oligomer conversion and expecting xylose conversions up to 90% (Humbird *et al.*, 2011).

Furthermore, studies on acid hydrolysis of EFB has been carried using H_2SO_4 as catalyst (Rahman *et al.*, 2007). After optimizing parameters such as temperature, reaction time and acid concentration, a xylose yield of 91% was obtained experimentally [(xylose obtained in liquor/maximum xylose concentration) x 100]. Considering 24% of xylan in EFB as reported in the study, this would already mean 240 g xyl/kg of EFB.

1.5. Alkali treatment

The basis of this treatment is the high solubilising effect bases have on lignin, when compared to cellulose and hemicellulose. It has been discussed that alkali treatment results in the saponification of ester bonds cross-binding xylan, hemicellulose and lignin (Misson *et al.*, 2009). Lime is the default base due to economic reasons, resulting in the biomass delignification with little hemicellulose hydrolysis (Alvira *et al.*, 2010). Besides improving the reactivity of remaining polysaccharides by removing lignin, some acetyl and uronic acid substitutions that lower enzyme accessibility are removed (Mosier *et al.*, 2005).

This pretreatment can be performed at ambient temperatures with a consequent increase in residence time that can be up to days of pretreatment. Treatment times can be lower by the use of high temperatures. To improve yields, addition of air or oxygen has been found positive for materials like poplar. However, for less lignified materials like corn stover, this approach presented only marginal improvements.

(Misson *et al.*, 2009) studied alkali pretreatment on EFB using 100mM NaOH and slake lime ($Ca(OH)_2$) for 48h, resulting in 65 and 9% lignin degradation. Improving the method with H_2O_2 consecutive treatment, resulted in complete delignification with NaOH and yields up to 44% using $Ca(OH)_2$. The oxidative characteristic of H_2O_2 might result in electrophilic substitution, displacement of side chains and cleavage of alkyl aryl ether linkages that improve lignin removal. However, as the oxidative reactions are not specific, this might result in lower sugar yields, with losses to side products. Furthermore, products derived from lignin oxidation might result in microbial growth inhibiting compounds that should be later removed (Hendriks *et al.*, 2009).

A study by (Rashid *et al.*, 2011) showed that NaOH treatment of EFB results in better saccharification (175 g of reducing sugars/kg of EFB, assuming a cellulose content of 36% this means 49 g/100g cellulose) when compared to H_2SO_4 and HNO_3, in the tested catalyst concentration range of 0.5 to 5%. If the improvement with H_2O_2 from the previous study is considered, this could result in promising yields, but no data was found in the literature.

On the other hand, $Ca(OH)_2$ treatment still has the benefit of being a low cost agent, with lower safety issues and can be easily recoverable as calcium carbonate by reaction with CO_2, with the possibility to converted back to lime with kiln technology (Alvira *et al.*, 2010).

Lignin recovery can be achieved by precipitation, which can be obtained by lowering the pH of the liquid fraction. In the study published by (Ibrahim *et al.*, 2004), it reported that H_3PO_4, H_2SO_4, HCl, and HNO_3 are suitable for recovering lignin from liquor resulting from EFB NaOH treatment. Of these acids, phosphoric and sulphuric acid yielded better results, 15.1 and 14.8 g lignin/L liqueur respectively at 20% acid concentrations. It was also found in this study that NaOH treated EFB lignin is similar to cereal straw and grass lignin.

1.6. Liquid hot water (LHW)

In this case, pressurised liquid water at temperatures of 160-240 °C is passed through a biomass bed, resulting in high conversion of hemicellulose and lignin to a lesser extent. Further cellulose conversion of the solid pretreated fraction is achieved at higher rate. This method is comparable to steam explosion, yielding a lightly higher pentosan recovery but in a more diluted form (Alvira *et al.*, 2010).

In the study of (Kim *et al.*, 2012), treatment of EFB was performed with both H_2SO_4 and LHW, the second one resulting in xylose and glucose yields about three times as low, ~40 g xylose/ kg EFB.

1.7. Ammonia fibre expansion (AFEX)

Similar to LHT, biomass is treated with liquid ammonia at high pressures. As the pressure is released the fibres in the biomass are expanded. Since the entering liquid ammonia leaves in gas form, the result is a solid stream with lower crystallinity cellulose and prehydrolysed hemicellulose.

One of the main advantages of AFEX is that there is no formation nof degradation products and thus, no conditioning is needed to neutralize or remove fermentation inhibitors (Alvira *et al.*, 2010).

The effects of AFEX pretreatment on EFB have been studied in different temperatures, finding yields (after saccharification) up to 210 g xyl/kg EFB and 162 g glc/kg EFB (57% glucose yield and 66% xylose yield) at 135 °C when biomass is put in contact with ammonia (1g/g NH_3) for 45 min (Lau *et al.*, 2010). Enzymes used also included xylanases and pectinases.

1.8. Organosolv

Organosolv as a pretreatment is similar to organosolv pulping, yielding high lignin removal with the use of organic solvents. As pretreatment it implies some benefits as easy solvent recovery by distillation and fractionation of lignocellulosic material into lignin, hemicellulose and cellulose fractions. Treatment with alcohols like ethanol and methanol has been proved successful; in contrast the use of organic acids is not recommended due to loss of solvent in acetylation of cellulose that might also reduce cellulose reactivity with enzymes.

Furthermore, organosolv also implies the major drawback of being very energy demanding in order to recover the solvent. Moreover, this pretreatment method implies high safety issues related to the volatility and toxicity of solvents. It has been considered too expensive to be used for biomass pretreatment (Zhao *et al.*, 2009).

2. Comparison of pretreatment technologies

Table A2. 1. Qualitative comparison of different pretreatment alternatives

Method	Advantages	Disadvantages
Mechanical comminution	Handles large sizes of biomass	Energy demanding
	Simple	Low sugar yield
	No inhibitors production	Requires additional steps
	Decreases cellulose crystallinity	
Biological	Degrades lignin	Low hydrolysis rate
	Degrades hemicellulose	
	Low energy requirements	
Steam explosion	Handles high solids loadings	Sugar degradation depending on severity
	Suitable for hard woods	Inhibitors production(requires conditioning)
	Suitable for agricultural residues	
	Hemicellulose solubilisation	Less effective on soft woods
	Lignin structure transformation	Partial xylan destruction
	Cost effective	Partial hemicellulose degradation
	Higher yields in two steps	
	Handles relatively large sizes	
	Feasible at industrial scale	
	No need of catalyst	
AFEX	Highly effective on agriculture residues and herbaceous crops	Hemicellulose is not hydrolysed (need of cellulase and hemicellulase)
	No sugar degradation	Ammonia capturing and recycle needed for economics
	Very high sugar yields in combination with enzymatic hydrolysis	Not efficient for biomass with high lignin content
	No inhibitors production	
	Increases accessible surface area	
	Removes lignin	
	Removes hemicellulose to some extent	
LHW	No need of catalyst	Energy demanding
	Hemicellulose solubilisation	Not feasible at large scale
	Increases accessible surface area	
Dilute acid	Widely studied and proven	Inhibitors production(requires conditioning)
	Effective for diverse feedstock	Sugar degradation depending on severity
	High sugar yields in combination with enzymatic hydrolysis	Low sugar concentration in exit stream
	Hemicellulose solubilisation	
	Alters lignin structure	
Lime	Hemicellulose solubilisation	Long residence time required
	Removes lignin	Irrecoverable salts incorporated to biomass
	Increases accessible surface area	Not very high sugar yields
Organosolv	Lignin solubilisation	Energy demanding
	Hemicellulose solubilisation	High safety risk compared to other methods
	Material fractionation	Solvents need to be recycled

3. Reviewed literature on main pretreatment alternatives

Table A2. 2. Reviewed literature for main pretreatments

Steam Explosion	Biomass	T (°C)	P (MPa)	Time (min)	Glc yield	Xyl yield	Total yield	Comments			Source
	EFB	140	0.28	60			31	Not accounting losses in condensate			Shamsudin, 2012
								Conditions available at mill			
	Grass straw	180	1.1	15	74	94	82	SL: 30%			Kumar, 2011
	Softwood	190	-	8	70	70	70	Best larger chip sizes			Ballesteros, 2002
		210	-	8	85	40	67	8-12mm			
	Corn stover	200	-	5	41	47	43				Zimbardi 2007
	Bagasse	205	-	10	67	44	58				Martin, 2002
Acid impregnated	Biomass	T (°C)	P (MPa)	Time (min)	Glc yield	Xyl yield	Total yield	Acid or Base	Load (%)	Comments	Source
	Corn stover	190		5	66	63	65	H2SO4	1.5		Zimbardi, 2007
		190			84	50	70	H2SO4	3		
	Bagasse	205	-	10	72	20	51	H2SO4	1	Higher inhibitors production	Martin, 2002
				10	71	54	64	SO2	1.1		
	Bagasse	190		5	92	82	88	SO2	2		Carrasco, 2010
Alkaline treatment	Grass straw	180	1.1	15	85	80	83	NaOH	1	Loading % w/wsolution	Kumar, 2011
	Corn stover	55		4 weeks	83	62	74			Non oxidative, N2 sparging	Kim, 2005
	Miscanthus	70		4	70	38	57		12	Extrusion	Vrije, 2002
Dilute acid	Grass straw	180			82	92	86	H2SO4	1	Sugars degradation is not accounted in yields	Kumar, 2011
	Corn stover	190			92	92	92	H2SO4	1.1		Kazi, 2010
	Switchgrass	190			81	97	88	H2SO4	1.1		Laser, 2009
	Corn stover	158	0.56	5	90	90	90	H2SO4	22.1 mg/g d biomass	Sugars degradation is not accounted in yields	Humingbird, 2011

APPENDIX 3. Process Alternatives

1. Process alternatives assumptions

1.1. Base Case

Lines crossing the battery limits are the inputs and outputs considered for the design project. Emissions are not accounted in this case as they are considered released to the environment. Steam calculations are based in the heat capacity of the biomass fuel, 13.76 MJ/kg on wet basis. Considering average utilization factors and heat to power ratios given by Husain *et al.* (2003) for CHP systems in palm oil mills, this results in 1 624 MJ/tonne FFB for heat and 91 MJ/tonne FFB for electricity. The reported energy requirements in mills vary greatly. Steam requirements range from 1 100 to 1 700 MJ/tonne FFB (Schmidt, 2007), with typical values around 1 400 MJ/tonne FFB. Thus, the available steam from shell and fibre combustion is considered sufficient. For electricity requirements, the typical reported values range from 52 to 80 MJ (15 to 22 kWh) per tonne FFB, with excess considered for administration, residence and laboratory buildings (Schmidt, 2007).

POME treatment emissions for lagoon systems had typically been calculated based on 0.21 - 0.25 kg CH_4/kg $COD_{removed}$ and 80% chemical oxygen demand (COD) removal efficiencies based on IPCC standards for wastewater (AES AgriVerde, 2008). In this report, recent measurements specifically for POME emissions on anaerobic lagoons are taken into account (Yacob *et al.*, 2006). Thus, methane emissions and COD removal efficiencies are taken as 0.235 kg CH_4/kg $COD_{removed}$ and 97.8% respectively.

1.2. Base Case Plus

Although current typical mills do not recover biogas from POME treatment, it is already considered for future mills. Therefore, as a complement to the Base Case, anaerobic POME treatment with biogas recovery is presented as Base Case Plus. The recovered amount of biogas is based on 95% COD removal efficiency achieved with semi-commercial closed anaerobic digester (Yacob *et al.*, 2006), assuming the same 0.235 kg CH_4/kg $COD_{removed}$ from the previous section. As biogas is considered for steam generation, it can be used to replace fibres and/or shells for fuel. Considering the same CPH utilization factors as from the previous section, the 23 GJ/h from biogas imply the replacement of 1.2 tonne/h of shells or 2.0 or tonne/h fibres. Considering moisture contents and biomass costs, it is assumed that biogas replaces 1.2 tonne/h shells in comparison to Base Case. This amount of shells is reduced from the battery-limit crossing shell stream, and thus a fraction of shells derived from palm oil milling is available for export.

1.3. Case A – Mill Derived Biomass

Case A comprises pretreatment and hydrolysis of the solid biomass waste derived from the oil palm mill operations. Fermentable sugars are to be obtained as the main product derived from biomass processing. Additionally, the use of the final solid residues, mainly as lignin, is intended for steam and power generation in combination with biogas obtained from POME anaerobic treatment.

Assumptions:

- Biomass is impregnated with 0.02 kg_{SO2}/kg dry biomass. 19% (w/w) of the acid is retained in the biomass as sulphuric and lignin sulfonic acids; the rest is available for recirculation.

- Steam requirements are based on required heat for bringing the biomass to 190 °C, considering calorific capacities of 2 and 4.18 MJ/tonne/°C of dry biomass and water respectively. It is considered that heat is available through saturated steam condensation at operating temperature.

- Based on enthalpy differences and evaporation enthalpy, it is taken that 17% (w/w) of steam is lost in the vent gas stream, along with volatile components.

- During pretreatment and enzymatic hydrolysis 92% and 84%(w/w) of the cellulose and hemicellulose present in the biomass are converted following the reactions given below, where 3.5%(w/w) of produced xylose is degraded to furfural, based on steam explosion results obtained in the study by (Carrasco *et al.*, 2010).

$(C_6H_{10}O_5)_n + n\ H_2O \longrightarrow n\ C_6H_{12}O_6$ Cellulose hydrolysis

$(C_5H_8O_4)_n + n\ H_2O \longrightarrow n\ C_5H_{10}O_5$ Hemicellulose hydrolysis

$C_5H_{10}O_5 \longrightarrow C_5H_4O_2 + 3\ H_2O$ Xylose degradation to furfural

- Enzymatic hydrolysis takes place at 20% solids loading, with 0.02 $kg_{protein}$/kg hollocellulose, based on NREL studies, (Humbird *et al.*, 2011).

- It is assumed that cellulose, hemicellulose and enzymes have the same heat of combustion, 17 350 MJ/tonne.
- The boiler efficiency and heat to power ratio are 65% of the total energy input and 18 to 1, as reported for several palm oil mills (Husain *et al.*, 2003).
- Steam at 190 °C and 140 °C is extracted from the combined heat and power generation system, for use during pretreatment and for export to the palm oil milling operations.
- Enzyme price is based on the cost contribution estimated in NREL studies with 63 RM/tonne of sugar.
- The price allocated to the fermentable sugar stream is based on the sugar content of the stream and the long-term contract price for sugar in Malaysia (see Appendix 1).
- Biogas recovery and COD removal efficiencies are 0.235 kg CH_4/kg $COD_{removed}$ and 97.8% respectively, as estimated in Base Case Plus with methane production.
- Both glucose and xylose are accounted for the fermentable sugars.

1.4. Case B – Mill and Plantation Biomass

In contrast to Case A, in Case B the biomass from both mill and plantation is taken into account. Thus, biomass streams are composed of OPF, OPT, EFB, shells and fibres. The rest remains the same as in Case A. Since OPF and OPT are not available at the mill, the collection and transportation cost are included in the cost of these biomass residues. Average transportation distance from plantation to mill is 5 km considering a circular plantation area of 18 000 ha, with the processing mill located at the centre.

Assumptions:

- Imbibition water use for OPT and OPF milling is 2 kg/kg of dry biomass.
- Sugar recovery from milling operation is 95% (w/w) of the total available sugar in the sap.

- Assumptions taken for pretreatment, enzymatic hydrolysis and cogeneration are the same as those mentioned for Case A.

- Glucose, xylose, fructose and sucrose (glucose + fructose) are accounted as fermentable sugars.

1.5. Case C – Multiple Mill Biomass

In this scenario biomass cost is affected, as the biomass has to be transported from several mills to the central one. The inputs and outputs are directly proportional to the number of mills and increase linearly, with exception of POME that is only available from the on-site mill. As result, the available biogas derived from POME treatment and thus the generated steam, are the same as for 1 mill. The prices are considered the same, except for the biomass, which increases with increased transportation distance.

1.6. Case D - Multiple Mills and Plantation Biomass

For this case, the mill and plantation biomass from several mills is taken into account, which affects directly the cost of all the biomass residues that have to be transported. The inputs and outputs are directly proportional to the number of mills and increase linearly, with exception of POME that is only available from the on-site mill. The prices are considered the same, except for the biomass, which increases with increased transportation distance.

2. Process alternatives - Scale effects

Table A3. 1 Biomass availability and sugar potential

Scale	Biomass				Fermentable sugars	
Number of mills	(tonne/y)		(d tonne/y)		(tonne/y)	
	M	M+P	M	M+P	M	M+P
1	151 200	551 200	74 970	194 970	42 252	136 680
2	302 400	1 102 400	149 940	389 940	84 504	273 360
3	453 600	1 653 600	224 910	584 910	126 756	410 040
4	604 800	2 204 800	299 880	779 880	169 008	546 720
5	756 000	2 756 000	374 850	974 850	211 260	683 400
6	907 200	3 307 200	449 820	1 169 820	253 512	820 080
7	1 058 400	3 858 400	524 790	1 364 790	295 764	956 760
8	1 209 600	4 409 600	599 760	1 559 760	338 016	1 093 440
9	1 360 800	4 960 800	674 730	1 754 730	380 268	1 230 120
10	1 512 000	5 512 000	749 700	1 949 700	422 520	1 366 800
11	1 663 200	6 063 200	824 670	2 144 670	464 772	1 503 480
12	1 814 400	6 614 400	899 640	2 339 640	507 024	1 640 160
13	1 965 600	7 165 600	974 610	2 534 610	549 276	1 776 840
14	2 116 800	7 716 800	1 049 580	2 729 580	591 528	1 913 520
15	2 268 000	8 268 000	1 124 550	2 924 550	633 780	2 050 200
16	2 419 200	8 819 200	1 199 520	3 119 520	676 032	2 186 880
17	2 570 400	9 370 400	1 274 490	3 314 490	718 284	2 323 560
18	2 721 600	9 921 600	1 349 460	3 509 460	760 536	2 460 240
19	2 872 800	10 472 800	1 424 430	3 704 430	802 788	2 596 920
20	3 024 000	11 024 000	1 499 400	3 899 400	845 040	2 733 600

Table A3. 2. Main economic indicators according to scale

Mills	Milling (million RM)		Pretreatment (million RM)		Enzymatic Hyd. (million RM)		CAPEX (million RM)		Capital charge (RM/tonne sugar)		TMC (RM/tonne sugar)	
	M	M+P	M	M+P	M	M+P	M	M+P	M	M+P	M	M+P
1	0.00	7.15	14.50	28.31	11.81	23.07	96	212	314.50	216.24	1293	1136
2	0.00	12.90	23.56	45.99	19.19	37.47	155	350	255.46	178.00	1225	1114
3	0.00	18.20	31.29	61.09	25.49	49.77	206	468	226.20	158.93	1206	1113
4	0.00	23.25	38.27	74.72	31.18	60.87	252	577	207.49	146.70	1201	1120
5	0.00	28.10	44.74	87.35	36.45	71.16	295	677	194.06	137.88	1202	1130
6	0.00	32.81	50.83	99.24	41.41	80.85	335	773	183.73	131.09	1206	1141
7	0.00	37.40	56.62	110.55	46.13	90.06	373	864	175.43	125.62	1211	1154
8	0.00	41.90	62.17	121.38	50.65	98.88	410	952	168.54	121.07	1218	1166
9	0.00	46.31	67.51	131.81	55.00	107.38	445	1036	162.69	117.20	1225	1179
10	0.00	50.65	72.68	141.90	59.21	115.60	479	1119	157.62	113.84	1232	1191
11	0.00	54.92	77.69	151.69	63.30	123.58	512	1199	153.18	110.90	1240	1203
12	0.00	59.14	82.57	161.22	67.27	131.34	544	1277	149.23	108.27	1249	1215
13	0.00	63.30	87.33	170.51	71.15	138.90	575	1353	145.69	105.92	1257	1227
14	0.00	67.42	91.98	179.59	74.94	146.30	606	1428	142.49	103.79	1265	1239
15	0.00	71.49	96.53	188.47	78.65	153.54	636	1501	139.57	101.84	1274	1250
16	0.00	75.52	101.00	197.18	82.28	160.64	665	1573	136.90	100.06	1282	1262
17	0.00	79.52	105.37	205.73	85.85	167.60	694	1644	134.43	98.41	1290	1273
18	0.00	83.48	109.68	214.13	89.35	174.44	722	1714	132.14	96.88	1298	1284
19	0.00	87.40	113.91	222.39	92.80	181.17	750	1782	130.02	95.46	1307	1294
20	0.00	91.30	118.07	230.52	96.19	187.79	778	1850	128.03	94.13	1315	1305

CAPEX: Capital expenditure, TMC: Total manufacturing costs.

Table A3. 3. Main economic indicators according to scale, in US dollars

Mills	Milling (million USD)		Pretreatment (million USD)		Enzymatic Hyd. (million USD)		CAPEX (million USD)		Capital charge (USD/tonne sugar)		TMC (USD/tonne sugar)	
	M	M+P	M	M+P	M	M+P	M	M+P	M	M+P	M	M+P
1	0.00	2.32	4.70	9.17	3.83	7.47	31	69	101.90	70.06	419	368
2	0.00	4.18	7.63	14.90	6.22	12.14	50	113	82.77	57.67	397	361
3	0.00	5.90	10.14	19.79	8.26	16.12	67	152	73.29	51.49	391	361
4	0.00	7.53	12.40	24.21	10.10	19.72	82	187	67.23	47.53	389	363
5	0.00	9.10	14.50	28.30	11.81	23.06	95	219	62.88	44.67	389	366
6	0.00	10.63	16.47	32.15	13.42	26.19	108	250	59.53	42.47	391	370
7	0.00	12.12	18.35	35.82	14.95	29.18	121	280	56.84	40.70	392	374
8	0.00	13.58	20.14	39.33	16.41	32.04	133	308	54.61	39.23	394	378
9	0.00	15.00	21.87	42.71	17.82	34.79	144	336	52.71	37.97	397	382
10	0.00	16.41	23.55	45.98	19.19	37.45	155	362	51.07	36.89	399	386
11	0.00	17.80	25.17	49.15	20.51	40.04	166	388	49.63	35.93	402	390
12	0.00	19.16	26.75	52.23	21.80	42.55	176	414	48.35	35.08	405	394
13	0.00	20.51	28.30	55.24	23.05	45.01	186	438	47.20	34.32	407	398
14	0.00	21.84	29.80	58.19	24.28	47.40	196	463	46.17	33.63	410	401
15	0.00	23.16	31.28	61.07	25.48	49.75	206	486	45.22	33.00	413	405
16	0.00	24.47	32.72	63.89	26.66	52.05	216	510	44.35	32.42	415	409
17	0.00	25.76	34.14	66.66	27.81	54.30	225	533	43.55	31.89	418	412
18	0.00	27.05	35.53	69.38	28.95	56.52	234	555	42.81	31.39	421	416
19	0.00	28.32	36.91	72.05	30.07	58.70	243	577	42.13	30.93	423	419
20	0.00	29.58	38.25	74.69	31.17	60.84	252	599	41.48	30.50	426	423

CAPEX: Capital expenditure, TMC: Total manufacturing costs.

Table A3. 4. Learning effect on main economic indicators for several scales

Mills	Milling (million RM)		Pretreatment (million RM)		Enzymatic Hyd. (million RM)		CAPEX (million RM)		Capital charge (RM/tonne sugar)		TMC (RM/tonne sugar)	
	M	M+P	M	M+P	M	M+P	M	M+P	M	M+P	M	M+P
1	0.00	7.15	4.97	9.71	4.05	7.91	33	90	107.87	91.54	871	888
2	0.00	12.90	8.08	15.78	6.58	12.85	53	151	87.62	76.70	889	915
3	0.00	18.20	10.73	20.95	8.74	17.07	71	204	77.59	69.24	911	937
4	0.00	23.25	13.13	25.63	10.69	20.88	86	253	71.17	64.42	931	959
5	0.00	28.10	15.35	29.96	12.50	24.41	101	299	66.56	60.93	950	979
6	0.00	32.81	17.43	34.04	14.20	27.73	115	343	63.02	58.24	968	999
7	0.00	37.40	19.42	37.92	15.82	30.89	128	386	60.17	56.06	985	1018
8	0.00	41.90	21.32	41.63	17.37	33.92	140	426	57.81	54.24	1000	1036
9	0.00	46.31	23.16	45.21	18.87	36.83	153	466	55.80	52.69	1015	1053
10	0.00	50.65	24.93	48.67	20.31	39.65	164	504	54.07	51.34	1030	1069
11	0.00	54.92	26.65	52.03	21.71	42.39	176	542	52.54	50.16	1043	1085
12	0.00	59.14	28.32	55.30	23.07	45.05	187	579	51.19	49.10	1057	1100
13	0.00	63.30	29.96	58.48	24.40	47.64	197	615	49.97	48.15	1070	1115
14	0.00	67.42	31.55	61.60	25.70	50.18	208	650	48.87	47.29	1082	1129
15	0.00	71.49	33.11	64.65	26.98	52.66	218	685	47.87	46.50	1094	1143
16	0.00	75.52	34.64	67.63	28.22	55.10	228	720	46.96	45.78	1106	1156
17	0.00	79.52	36.14	70.57	29.45	57.49	238	753	46.11	45.11	1118	1169
18	0.00	83.48	37.62	73.45	30.65	59.83	248	787	45.32	44.49	1129	1182
19	0.00	87.40	39.07	76.28	31.83	62.14	257	820	44.60	43.91	1140	1194
20	0.00	91.30	40.50	79.07	32.99	64.41	267	852	43.91	43.37	1151	1206

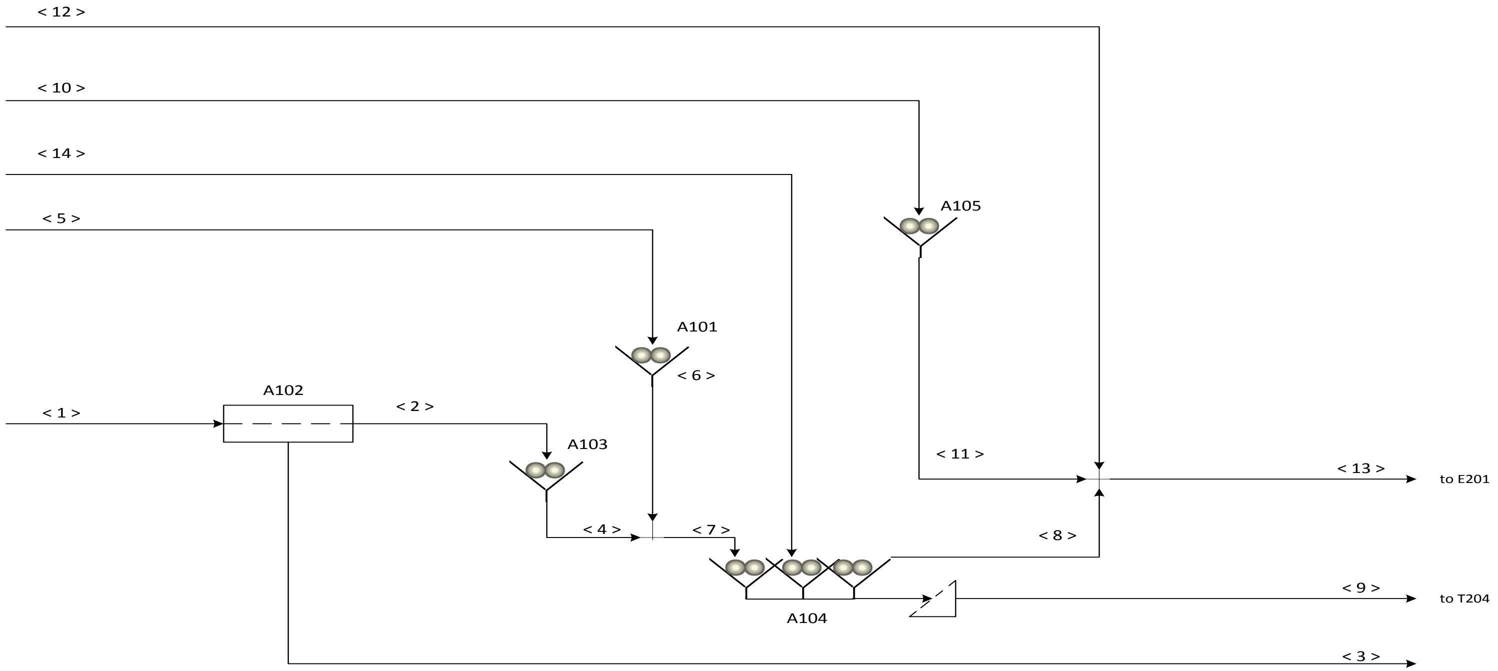
CAPEX: Capital expenditure, TMC: Total manufacturing costs.

Table A3. 5. Learning curve effects on main economic indicators for several scales, in US dollars

Mills	Milling (million USD)		Pretreatment (million USD)		Enzymatic Hyd. (million USD)		CAPEX (million USD)		Capital charge (USD/tonne sugar)		TMC (USD/tonne sugar)	
	M	M+P	M	M+P	M	M+P	M	M+P	M	M+P	M	M+P
1	0.00	2.32	1.61	3.15	1.31	2.56	11	29	34.95	29.66	282	288
2	0.00	4.18	2.62	5.11	2.13	4.16	17	49	28.39	24.85	288	296
3	0.00	5.90	3.48	6.79	2.83	5.53	23	66	25.14	22.43	295	303
4	0.00	7.53	4.25	8.30	3.46	6.76	28	82	23.06	20.87	302	311
5	0.00	9.10	4.97	9.71	4.05	7.91	33	97	21.57	19.74	308	317
6	0.00	10.63	5.65	11.03	4.60	8.98	37	111	20.42	18.87	314	324
7	0.00	12.12	6.29	12.29	5.13	10.01	41	125	19.50	18.16	319	330
8	0.00	13.58	6.91	13.49	5.63	10.99	46	138	18.73	17.57	324	336
9	0.00	15.00	7.50	14.65	6.11	11.93	49	151	18.08	17.07	329	341
10	0.00	16.41	8.08	15.77	6.58	12.85	53	163	17.52	16.63	334	346
11	0.00	17.80	8.63	16.86	7.03	13.73	57	176	17.02	16.25	338	351
12	0.00	19.16	9.18	17.92	7.48	14.60	60	188	16.58	15.91	342	356
13	0.00	20.51	9.71	18.95	7.91	15.44	64	199	16.19	15.60	347	361
14	0.00	21.84	10.22	19.96	8.33	16.26	67	211	15.84	15.32	351	366
15	0.00	23.16	10.73	20.95	8.74	17.06	71	222	15.51	15.07	355	370
16	0.00	24.47	11.22	21.91	9.14	17.85	74	233	15.21	14.83	358	375
17	0.00	25.76	11.71	22.86	9.54	18.63	77	244	14.94	14.62	362	379
18	0.00	27.05	12.19	23.80	9.93	19.39	80	255	14.69	14.41	366	383
19	0.00	28.32	12.66	24.71	10.31	20.13	83	266	14.45	14.23	369	387
20	0.00	29.58	13.12	25.62	10.69	20.87	86	276	14.23	14.05	373	391

CAPEX: Capital expenditure, TMC: Total manufacturing costs.

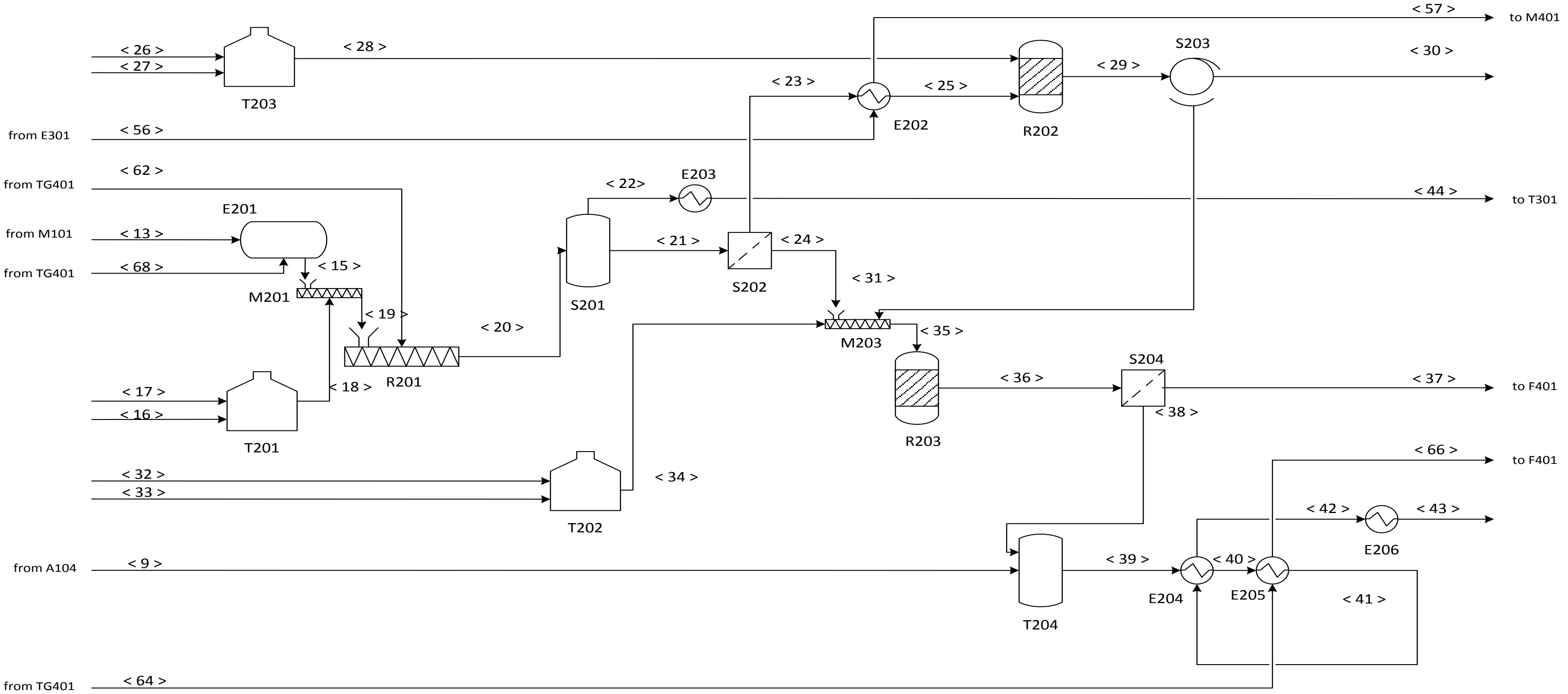
APPENDIX 4. Process Flow Diagrams



A 100 - STREAMS		
No.	ASPEN CODE	Stream
1	OPT	Oil palm trunk
2	OPTA	Debarked Oil Palm trunk
3	BARK	Bark
4	SOPT	Shredded oil palm trunk
5	OPF	Oil palm frond
6	SOPF	Shredded oil palm frond
7	PBM	Shredded plantation biomass
8	PBMF	Shredded plantation biomass fibres
9	SAP	Sap
10	EFB	Empty fruit bunches
11	SEFB	Shredded empty fruit bunches
12	FIBRES	Mesocarpic fibre
13	SBM	Shredded biomass
14	IMB WATER	Imbibition water

A 100 - EQUIPMENT	
Code	Equipment
A101	Frond Shredder
A102	Trunk debarker
A103	Trunk shredder
A104	Mill tandem
A105	EFB shredder

Mill Integrated Conversion of Oil Palm Biomass into Commodity Chemical Intermediates
A 100 - BIOMASS PRE-PROCESSING
M. D. M. Palmeros Parada
PDEng Trainee
TU Delft - OPBC
December, 2012

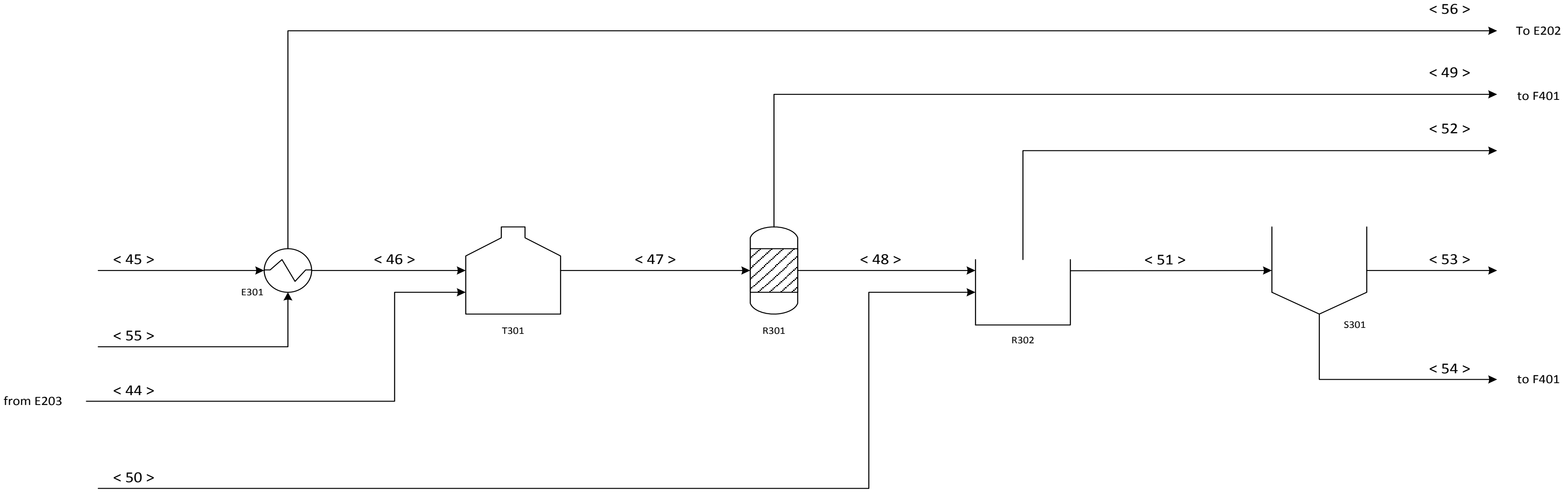


A 200 - STREAMS					
No.	ASPEN CODE	Stream	No.	ASPEN CODE	Stream
15	HBM	Preheated biomass	31	CONDLIQ	Conditioned liquid
16	SACID	Concentrated acid	32	ENZ	Enzyme
17	DILWATER	Dilution water for acid	33	DILWAT2	Dilution water for enzyme
18	SACIDSOL	Diluted acid	34	ENZSLN	Enzyme solution
19	AIBM	Acid impregnated biomass	35	PREENZ	Slurry for enzymatic hydrolysis
20	PTTBM	Pretreated biomass	36	ENZSLUR	Hydrolysate slurry
21	PTTBM2	Flashed pretreated biomass	37	LIGSOLID	Ligneous solids
22	VENTA	Flash vent	38	LCSUG	LC sugars
23	PTTLIQ	Pretreated liquid	39	SUG	Sugars
24	PTTSOLID	Pretreated solids	40	S1	Sugars
25	CPTTLIQ	Cold pretreated solids	41	S2	Sugars
26	SLIME	Solid Lime	42	S3	Sugars
27	WAT3	Dilution water for lime	43	PSUG	Sugar product
28	LIME	Lime solution	44	VENTB	Condensed vent
29	CONDSLUR	Conditioned slurry	67	STMCOND	Condensed steam
30	GYPSUM	Gypsum			

A 200 - EQUIPMENT			
Code	Equipment	Code	Equipment
E201	Presteamer	R203	Enzymatic hydrolysis reactor
M201	Screw feeder	S204	Screw press
R201	Pretreatment reactor	T201	Acid tank
S201	Flash tank	T202	Enzyme tank
E203	Vent condenser	T203	Lime tank
S202	Screw press	T204	Sugars tank
E202	Pttliq cooler	E204	Sugars heater 1
R202	Conditioning reactor	E205	Sugars heater 2
S203	Drum filter	E206	Sugars cooler
M203	Screw feeder		

Mill Integrated Conversion of Oil Palm Biomass
into Commodity Chemical Intermediates

A 200 - BIOMASS CONVERSION
M. D. M. Palmeros Parada
PDEng Trainee
TU Delft - OPBC
December, 2012

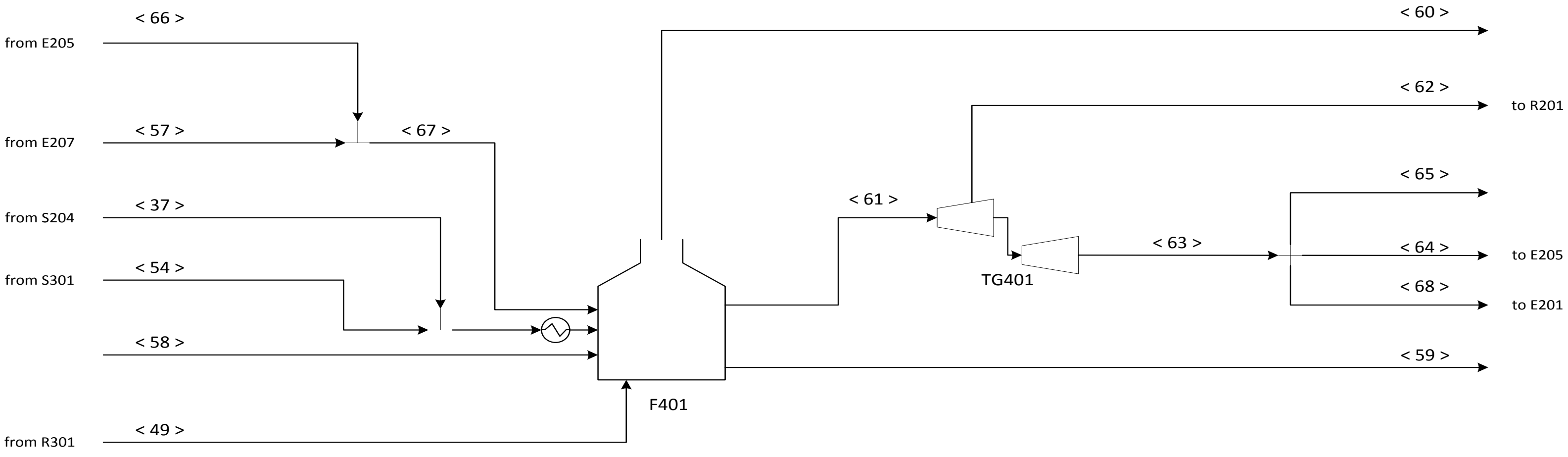


A 300 - STREAMS		
No.	ASPEN CODE	Stream
45	POME	Palm oil mill effluent
46	POMEB	Cooled palm oil mill effluent
47	WWAT	Waste water
48	TTWAT	Treated water
49	BIOGAS	Biogas
50	LAGAIR	Air from environment
51	TTWATC	Treated water
52	LOSSES	Losses into environment
53	TTPOME	Treated water
54	SLUDGE2	Sludge

A 300 - EQUIPMENT	
Code	Equipment
T301	Waste water tank
E301	POME cooler
R301	Anaerobic reactor
R302	Aerobic Basin
S301	Clarifier

Mill Integrated Conversion of Oil Palm Biomass into Commodity Chemical Intermediates

A 300 - WASTE WATER TREATMENT
M. D. M. Palmeros Parada
PDEng Trainee
TU Delft - OPBC
December, 2012



A 400 - POWER AND HEAT GENERATION		
No.	ASPEN CODE	Stream
55	BFW	Boiler feed water
56	BFWB	Preheated boiler feed water
57	BFWC	Preheated boiler feed water
58	COMBAIR	Combustion air
59	COMBSOL	Combustion solids
60	FLUEG	Flue gas
61	SHSM	Super-heated steam
62	MPSTMA	Medium Pressure steam
63	LPSTM	Low pressure team
64	LPSTEAM	Low pressure team
65	LPSTEAM	Low pressure team
66	STMCOND	Condensed steam
67	BFWD	Preheated boiler feed water
68	LPSTEAM	Low pressure team

A 400 - EQUIPMENT	
Code	Equipment
M401	Mixer
F401	Furnace
TG401	Turbine, extraction point 1
TG401B	Turbine, extraction point 2

Mill Integrated Conversion of Oil Palm Biomass into Commodity Chemical Intermediates

A 400 - POWER AND HEAT GENERATION
M. D. M. Palmeros Parada
PDEng Trainee
TU Delft - OPBC
December, 2012

APPENDIX 5. Mass and Heat Balance Tables

A5. 1 Mass and Heat Balance - Three-mill Case

IN						OUT				
Plant		EQUIPMENT			EQUIPM.	EQUIPMENT			Plant	
Mass tonne/h	Heat kW	Mass tonne/h	Heat kW	Stream Nr.	IDENTIF.	Stream Nr.	Mass tonne/h	Heat kW	Mass tonne/h	Heat kW
110	-390147	110	-390147	< 5 >	A101	< 6 >	110.0	-390147		
110	**-390147**	**110**	**-390147**		**Total**		**110.0**	**-390147**		
90	-317673	90	-317673	< 1 >	A102	< 2 >	78.3	-276376		
						< 3 >	11.7	-41298	12	-41298
90	**-317673**	90	**-317673**		Total		90.0	**-317673**	**12**	**-41298**
		78	-276376	< 2 >	A103	< 4 >	78.3	-276376		
		78	**-276376**		**Total**		**78.3**	**-276376**		
		110	-390147	< 6 >	A104	< 8 >	94.6	-277111		
		78	-276376	< 4 >		< 9 >	178.4	-762635		
85	-373224	85	-373224	< 14 >						
85	**-373224**	**273**	**-1039747**		**Total**		**273.0**	**-1039747**		
41	-142867	41	-142867	< 10 >	A105	< 11 >	41.4	-142867		
41	**-142867**	**41**	**-142867**		**Total**		**41.4**	**-142867**		
		95	-277111	< 8 >	M101	< 13 >	160.3	-481393		
		41	-142867	< 11 >						
24	-61415	24	-61415	< 12 >						
24	**-61415**	**160**	**-481393**		**Total**		**160.3**	**-481393**		
		160	-481393	< 13 >	E201	< 15 >	171.7	-523305		
		11	-41913	< 68 >						
		172	**-523305**		**Total**		**171.7**	**-523305**		
		172	-523305	< 15 >	R201	< 20 >	302.0	-1052678		
		78	-340645	< 18 >						
		52	-188728	< 62 >						
		302	**-1052678**		**Total**		**302.0**	**-1052678**		
		302	-1052678	< 20 >	S201	< 21 >	253.4	-871378		
						< 22 >	49.0	-181300		
		302	**-1052678**		**Total**		**302.4**	**-1052678**		
		147	-587026	< 23 >	E202	< 25 >	147.4	-595874		
		130	-569876	< 56 >		< 57 >	130.0	-561031		
		277	**-1156902**		**Total**		**277.4**	**-1156905**		
		49	-181300.2	< 22 >	E203	< 44 >	49.0	-215634		
	-34334		-34334	CW						
	-34334		**-215634**		Total			**-215634**		
		253	-871378	< 21 >	S202	< 23 >	147.4	-587026		
						< 24 >	106.0	-284352		
		253	**-871378**		**Total**		**253.4**	**-871378**		
		147	-595874	< 25 >	R202	< 29 >	159.0	**-648191**		
		12	-52317	< 28 >						
		159	**-648191**		**Total**		**159.0**	**-648191**		
		159	-648191	< 29 >	S203	< 31 >	157.0	-640989		
						< 30 >	2.0	-7202	2	-7202
		159	**-648191**		**Total**		**159.0**	**-648191**	**2**	**-7202**
		157	-640989	< 31 >	M203	< 35 >	382.0	-1448173		

		106	-284352	< 24 >						
		119	-522832	< 34 >						
		382	**-1448173**		**Total**		**382.0**	**-1448173**		
		382	-1448173	< 35 >	R203	< 36 >	382.0	-1448173		
		382	**-1448173**		**Total**		**382.0**	**-1448173**		
		382	-1448173	< 36 >	S204	< 37 >	41.6	-94641		
						< 38 >	340.0	-1353532		
		382	**-1448173**		**Total**		**381.6**	**-1448173**		
2	-3883	2	-3883	< 16 >	T201	< 18 >	78.0	-340645		
76	-336762	76	-336762	< 17 >						
78	**-340645**	**78**	**-340645**		**Total**		**78.0**	**-340645**		
0	-21	0	-21	< 32 >	T202	< 34 >	118.8	-522832		
119	-522811	119	-522811	< 33 >						
119	**-522832**	**119**	**-522832**		**Total**		**118.8**	**-522832**		
1	-4463	1	-4463	< 26 >	T203	< 28 >	12.1	-52317		
11	-47854	11	-47854	< 27 >						
12	**-52317**	**12**	**-52317**				**12.1**	**-52317**		
		178	-762635	< 9 >	T204	< 39 >	518.4	-2116167		
		340	-1353532	< 38 >						
		518	**-2116167**		**Total**		**518.4**	**-2116167**		
		518	-2116167	< 39 >	E204	< 40 >	518.4	-2094243		
		518	-2091196	< 41 >		< 42 >	518.0	-2113120		
		1036	**-4207363**		**Total**		**1036.4**	**-4207363**		
		518	-2094243	< 40 >	E205	< 41 >	518.0	-2091196		
		5	-17087	< 64 >		< 66 >	4.7	-20133		
		523	**-2111330**		**Total**		**522.7**	**-2111329**		
		518	-2113120	< 42 >	E206	< 43 >	518.0	-2124244	518	-2124244
	-11124		-11124	CW						
	-11124	**518**	**-2124244**		**Total**		**518.0**	**-2124244**	**518**	**-2124244**
40	-171469	40	-171469	< 45 >	E301	< 46 >	40.4	-173280		
130	-571392	130	-571392	< 55 >		< 56 >	130.0	-569876		
	-295		-295	CW						
170	**-743156**	**170**	**-743156**		**Total**		**40.4**	**-743156**		
		40	-173277	< 46 >	T301	< 47 >	89.8	-388912		
		49	-215634	< 44 >						
		90	**-388912**		**Total**		**89.8**	**-388912**		
		90	-388912	< 47 >	R301	< 48 >	87.8	-384514		
						< 49 >	2.0	-4398		
		90	**-388912**		**Total**		**89.8**	**-388912**		
		88	-384514	< 48 >	R302	< 51 >	87.8	-386562		
10	-1099	10	-1099	< 50 >		< 52 >	10.0	-1408	10	-1408
						Heat to atm		2358		2358
10	**-1099**	**98**	**-385613**		**Total**		**97.8**	**-385613**	**10**	**949**
		88	-386562	< 51 >	S301	< 53 >	86.9	-382244	87	-382244
						< 54 >	0.9	-4318		
		88	**-386562**		**Total**		**87.8**	**-386562**	**87**	**-382244**
		130	-561031	< 57 >	M401	< 67 >	134.0	-581164		
		5	-20133	< 66 >						
		134	**-581164**		**Total**		**134.0**	**-581164**		
		134	-581164	< 67 >	F401	< 59 >	4.7	-13801	5	-13801
		1	-4318	< 54 >		< 60 >	243.6	-251487	244	-251487
		42	-94641	< 37 >		< 61 >	134.0	-469518		

		2	-4398	< 49 >		Eff. Losses		27911		27911
204	-22374	204	-22374	< 58 >						
204	**-22374**	**382**	**-706894**		**Total**		**382.3**	**-706894**	**248**	**-237376**
		134	-469518	< 61 >	TG401	< 62 >	52.0	-188728		
			-19416.4	POWER		< 63 >	82.0	-300206		
		134	**-488934**		**Total**		**134.0**	**-488934**		
		82	-300210	< 63 >	SPLIT	< 68 >	11.4	-41913		
						< 64 >	5.0	-17087		
						< 65 >	66.0	-241210	66	-241210
		82	**-300210**		**Total**		**82.4**	**-300210**	**66**	**-241210**
943	**-3013206**				Total				**943**	**-3032624**
OUT - IN (%) :									0.0%	0.6%

A5. 2. Ten-mill Case - Mass and Heat Balances

IN						OUT				
Plant		EQUIPMENT			EQUIPM.	EQUIPMENT			Plant	
Mass tonne/h	Heat kW	Mass tonne/h	Heat kW	Stream Nr.	IDENTIF.	Stream Nr.	Mass tonne/h	Heat kW	Mass tonne/h	Heat kW
367	-1300492	367	-1300492	< 5 >	A101	< 6 >	367	-1300492		
367	**-1300492**	**367**	**-1300492**		**Total**		**367**	**-1300492**		
300	-1058901	300	-1058901	< 1 >	A102	< 2 >	261	-921244		
						< 3 >	39	-137657	39	-137657
300	**-1058901**	**300**	**-1058901**		**Total**		**300**	**-1058901**	**39**	**-137657**
		261	-921244	< 2 >	A103	< 4 >	261	-921244		
		261	**-921244**		**Total**		**261**	**-921244**		
		366	-1300492	< 6 >	A104	< 8 >	315	-923799		
		261	-921244	< 4 >		< 9 >	595	-2542052		
282	-1244115	282	-1244115	< 14 >						
282	**-1244115**	**910**	**-3465851**		**Total**		**910**	**-3465851**		
138	-475907	138	-475907	< 10 >	A105	< 11 >	138	-475907		
138	**-475907**	**138**	**-475907**		**Total**		**138**	**-475907**		
		315	-923799	< 8 >	M101	< 13 >	534	-1603944		
		138	-475907	< 11 >						
81	-204238	81	-204238	< 12 >						
81	**-204238**	**534**	**-1603944**		**Total**		**534**	**-1603944**		
		534	-1603944	< 13 >	E201	< 15 >	572	-1743584		
		38	-139640	< 68 >						
		572	**-1743584**		**Total**		**572**	**-1743584**		
		572	-1743584	< 15 >	R201	< 20 >	1011	-3519851		
		263	-1144374	< 18 >						
		176	-631895	< 62 >						
		1011	**-3519853**		**Total**		**1011**	**-3519851**		
		1011	-3519851	< 20 >	S201	< 21 >	845	-2913543		
						< 22 >	165	-606308		
		1011	**-3519851**		**Total**		**1011**	**-3519851**		
		492	-1963564	< 23 >	E202	< 25 >	492	-1993188		
		413	-1815954	< 56 >		< 57 >	413	-1786330		
		904	**-3779518**		**Total**		**904**	**-3779518**		
		165	-606308	< 22 >	E203	< 44 >	165	-721157		
	-114849		-114849	CW						
	-114849	**165**	**-721157**		**Total**		**165**	**-721157**		
		845	-2913543	< 21 >	S202	< 23 >	492	-1963564		
						< 24 >	354	-949979		
		845	**-2913543**		**Total**		**845**	**-2913543**		
		492	-1993188	< 25 >	R202	< 29 >	530	-2158527		
		38	-165339	< 28 >						
		530	**-2158527**		**Total**		**530**	**-2158527**		
		530	-2158527	< 29 >	S203	< 31 >	522	-2134535		
						< 30 >	7	-23991	7	-23991
		530	**-2158527**		**Total**		**530**	**-2158526**	**7**	**-23991**
		522	-2134535	< 31 >	M203	< 35 >	1274	-4836614		
		354	-949979	< 24 >						
		398	-1752100	< 34 >						
		1274	**-4836614**		**Total**		**1274**	**-4836614**		

		1274	-4836614	< 35 >	R203	< 36 >	1274	-4836614		
		1274	**-4836614**		**Total**		**1274**	**-4836614**		
		1274	-4836614	< 36 >	S204	< 37 >	139	-315396		
						< 38 >	1135	-4521219		
		1274	**-4836614**		**Total**		**1274**	**-4836615**		
6	-12932	6	-12932	< 16 >	T201	< 18 >	263	-1144374		
257	-1131441	257	-1131441	< 17 >						
263	**-1144373**	**263**	**-1144373**		**Total**		**263**	**-1144374**		
0	-21	0	-21	< 32 >	T202	< 34 >	398	-1752100		
398	-1752079	398	-1752079	< 33 >						
398	**-1752100**	**398**	**-1752100**		**Total**		**398**	**-1752100**		
4	-14103	4	-14103	< 26 >	T203	< 28 >	38	-165338		
34	-151235	34	-151235	< 27 >						
38	**-165338**	**38**	**-165338**				**38**	**-165338**		
		595	-2542052	< 9 >	T204	< 39 >	1730	-7063270		
		1135	-4521219	< 38 >						
		1730	**-7063271**		**Total**		**1730**	**-7063270**		
		1730	-7063270	< 39 >	E204	< 40 >	1730	-6990080		
		1730	-6979933	< 41 >		< 42 >	1730	-7053123		
		3460	**-14043203**		**Total**		**3460**	**-14043203**		
		1730	-6990080	< 40 >	E205	< 41 >	1730	-6979933		
		16	-56903	< 64 >		< 66 >	16	-67051		
		1745	**-7046983**		**Total**		**1745**	**-7046984**		
		1730	-7053123	< 42 >	E206	< 43 >	1730	-7090262	1730	-7090262
	-37138		-37138	CW						
	-37138	**1730**	**-7090261**		**Total**		**1730**	**-7090262**	**1730**	**-7090262**
40	-171469	40	-171469	< 45 >	E301	< 46 >	40	-173277		
413	-1817762	413	-1817762	< 55 >		< 56 >	413	-1815954		
453	**-1989231**	**453**	**-1989231**		**Total**		**453**	**-1989231**		
		40	-173277	< 46 >	T301	< 47 >	206	-894435		
		165	-721157	< 44 >						
		206	**-894434**		**Total**		**206**	**-894435**		
		206	-894435	< 47 >	R301	< 48 >	203	-887758		
						< 49 >	3	-6676		
		206	**-894435**		**Total**		**206**	**-894434**		
		203	-887758	< 48 >	R302	< 51 >	203	-891777		
10	-1099	10	-1099	< 50 >		< 52 >	10	-1826	10	-1826
						Heat to atm		4746		4746
10	**-1099**	**213**	**-888856**		**Total**		**213**	**-888857**	**10**	**2920**
		203	-891777	< 51 >	S301	< 53 >	200	-882386	200	-882386
						< 54 >	2	-9391		
		203	**-891777**		**Total**		**203**	**-891777**	**200**	**-882386**
		413	-1786330	< 57 >	M401	< 67 >	428	-1853381		
		16	-67051	< 66 >						
		428	**-1853381**		**Total**		**428**	**-1853381**		
		428	-1853381	< 67 >	F401	< 59 >	16	-46010	16	-46010
		2	-9391	< 54 >		< 60 >	817	-807738	817	-807738
		139	-315396	< 37 >		< 61 >	428	-1496098		
		3	-6676	< 49 >		Eff. Losses		89321		89321
689	-75682	689	-75682	< 58 >						
689	**-75682**	**1261**	**-2260526**		**Total**		**1261**	**-2260525**	**833**	**-764427**

		428	-1496098	< 61 >	TG401	< 62 >	176	-631895		
			-61265.4	POWER		< 63 >	252	-925470		
		428	**-1557363**		**Total**		**428**	**-1557365**		
		252	-925470	< 63 >	SPLIT	< 68 >	38	-139640		
						< 64 >	16	-56903		
						< 65 >	199	-728927	199	-728927
		252	**-925470**		**Total**		**252**	**-925470**	**199**	**-728927**
3019	**-9563464**				Total				**3018**	**-9624730**
OUT - IN :									0.0%	0.6%

APPENDIX 6. Process and Equipment Design Calculations

1. Three mill cluster

1.1. OPF preparation

Typical dimensions for large capacity debarker units are presented in Table A6. 1, with information from Zhengzhou Yijiayi Fashion Industry Co., Ltd. as supplier (Alibaba.com, 2012).

Table A6. 1 Debarking equipment specifications (Alibaba.com, 2012).

Specifications	6 meter roller	9 meter roller	12 meter roller
Capacity (tonne/h)	7-15	15-25	25-30
Motor power (kW)	7.5*2	7.5*2	7.5*2
Length (mm)	6300	9000	12600
Width (mm)	1310	1500	1550
Height (mm)	1550	1700	1650
Weight (tonne)	4	5	8

1.2. OPF and OPT milling

In order to define the number of milling tandems necessary for processing the OPF and OPT streams, it is necessary to first establish the capacity of each milling tandem. According to Hugot (1986) the capacity of the mill can be calculated from equation A6.1.

$$A_c = \frac{0.9\,c\cdot n\cdot(1 - 0.06\,n\cdot D)\cdot L\cdot D^2\cdot\sqrt{N}}{f} \qquad \text{(equation A6.1)}$$

Where:

A_c = Capacity of the tandem, in tonne per hour

f = fibre fraction, as tonne of fibre/tonne

c = coefficient for preparatory plant, 1.22 for 2 knife sets and one shredder

L = length of rollers, in m

D = diameter of rollers, in m

n = speed of rotation of rollers, in rpm

N = number of rollers in the tandem

For roller dimensions, the standard values 762 x 1 524 mm are considered. Roller speeds are in the range of 6 rpm and number of rollers ranges from 9 – 21 rollers (3 – 7 sets). The most typical milling configurations are 1 crusher in combination with 4 and 5 mills (14 and 17 rollers respectively). Using the rollers speed of 6rpm, the mentioned typical configurations result in capacities of 53 and 58 tph, thus implying the need of 4 tandems in either case for covering the 188.3 tonne/h OPF and OPT stream. It is therefore considered a better option to install 4 tandems of 14 rollers (1 crusher plus 4 mill sets), that could process up to 211 tonne/hour. To have an estimation of the power required for each milling tandems Parr's formula can be used, equation A6.2.

$$P := 0.3f \cdot N^{0.45} \qquad \text{(equation A6.2)}$$

Where:
P = power absorbed by a tandem, in hp/tph
F = fibre %, as tonne of fibre per 100 tonne
N = number of rollers in the tandem

For the 4 tandems the power requirement is 4144 kW, which is in the typical range reported by Hugot (1986) for tandems of 14 or 15 rollers, 20 – 25 HP/tfh/mill.

In order to reach high recoveries it is a common practice to include imbibition in the milling operations. Sugar recovery is higher as more imbibition water is used, however the recovered sugars are in diluted form. Estimations for achieved sugar recovery in function of added water can be calculated using equations A6.3 and A6.4.

$$e := \frac{\alpha \cdot (mf - f)}{mf \cdot (1 - f)} + \left[1 - \frac{\alpha \cdot (mf - f)}{mf \cdot (1 - f)}\right] \cdot \lambda \cdot \frac{\lambda^{n} - 1}{\lambda^{(n+1)} - 1} \qquad \text{(equation A6.3)}$$

$$\lambda := \frac{\beta \cdot r}{1 - \beta \cdot r} \qquad \text{(equation A6.4)}$$

Where:
e = efficiency, as a fraction of recovered sugar
mf = remaining moisture in milled biomass
f = fibre fraction in biomass
α = coefficient for differences in purity of sugar in juice and milled biomass
λ = mass of added water per mass of fibres in the biomass
β = coefficient for fitting analytical to operational results

In practice, the purity of sugar is higher in the juice than in the remaining biomass, thus the coefficient α is typically 1.05 – 1.10. The coefficient β is used to fit the theoretical values to reported operational results, and in the range of 0.6 – 0.8. In the present study, 1.075 is sued for α and 0.8 is used for β.

Typical sugar recoveries for sugarcane milling are in the range of 95%, with reported imbibition water values in the range of 1.5 – 2.5 tonne of water per tonne of fibres. Using equation A6.3, it is estimated that with water addition at 2 tonne/tonne, 95% of the sugars are recovered, at a concentration of 4.5% (w/w). It is thus decided to use imbibition water as 1.5 tonne/tonne. The estimated sugar recovery is 93% and the sugar concentration is 5.3% under this imbibition regime. Optimum water used can be estimated in future works based on process economics, considering the amount of energy needed to process a more diluted product, versus the higher sugar recovery.

1.3. Pretreatment

Sulphuric acid is probably the most common acid catalyst for pretreatment; it is well known and relatively inexpensive. However, there are some disadvantages related to it when compared to SO_2, e.g. it is more corrosive. On the other hand, SO_2 use and recovery might have high cost contributions to the overall process, besides raising the process complexity by adding extra steps to the process. A techno-economic study showed that SO_2 treatment resulted more costly than for instance, dilute acid pretreatment due to SO_2 recovery (Schell *et al.*, 1991). Although this study could result different with conversion yields achievable today, it is considered as an indication in favour H_2SO_4 selection as catalyst. Additionally, sulphur dioxide is highly toxic and a strong air pollutant being the precursor of acid rain. Considering the strong environmental critiques to which oil palm production is already subjected to, it is considered to continue working with H_2SO_4. A literature review of yields at optimum conditions according to different studies is presented in Table A6. 2.

Based reported data, see Table A6. 3, it is clear that optimum temperatures for agricultural residues are typically in the range of 190 – 205 °C. Considering that a slight increase above optimum temperatures results in product degradation into furans, it is decided to work in the lower side of reported optimum temperatures, at 190 °C. Following the results obtained by Tucker et al. (2003) with corn-stover, it is considered that the pretreatment will be held for 2 minutes.

Table A6. 2 Hydrolysis yields after pretreatment with acid catalysts

Biomass	Temp (C)	Glc yield* (%)	Xyl yield* (%)	Acid Catalyst	Source
Corn stover	190	66	63	H_2SO_4	Zimbardi, 2007
Corn stover	190	84	50	H_2SO_4	Zimbardi, 2007
Bagasse	205	72	20	H_2SO_4	Martin, 2002
Willow chips	190	67	90	H_2SO_4	Eklund, 1995
Corn stover	190	90	90	H_2SO_4	Tucker, 2003
Bagasse		71	54	SO_2	Martin, 2002
Bagasse	190	92	82	SO_2	Carrasco, 2010
Aspen chips	-	-	75	SO_2	Schell, 1991
Switchgrass	195	92-94	-	SO_2	Ewanck, 2011
Sugarcane	205	92-94	-	SO_2	Ewanck, 2011
Willow chips	200	95	62	SO_2	Eklund, 1995
Corn fibre	190	95	95	SO_2	Mabee, 2006

*Glucose (Glc) and Xylose (Xyl) yields are percentages of theoretical yields after pretreatment and enzymatic hydrolysis.

Table A6. 3. Hydrolysis yields with different pretreatment conditions

Biomass	Time (min)	Glc yield* (%)	Xyl yield* (%)	Acid Catalyst (%**)		Solids (%)	Source
Corn stover	5	66	63	H_2SO_4	1.5	30-35	Zimbardi, 2007
Corn stover		84	50	H_2SO_4	3	30-35	Zimbardi, 2007
Bagasse	10	72	20	H_2SO_4	1	-	Martin, 2002
Willow chips		67	90	H_2SO_4	3	-	Eklund, 1995
Corn stover	1.5-2.0	>90	>90	H_2SO_4	1	45-55	Tucker, 2003

$^+$Yields are % of theoretical mass conversion yields, obtained after pretreatment and enzymatic hydrolysis.

**Catalyst loadings are reported under different basis, typically as the acid concentration of the injected solution, or the acid concentration in the liquid separated from the slurry. The values in the table are as given by the authors and are thus, not normalized.

Acid pretreatment equipment with and without steam explosion, is found already at pilot scale. There are also some patented designs for large scale by companies like Andritz Inc. Pretreatment reactors operating vertically have been proposed based on paper and pulp industry experience. However, it has been observed that horizontal operation allows tighter residence times distributions, which is highly important for high severity single-step pretreatments as proposed in the present project. The 2011 NREL design, with a similar set-up to the Andritz steam explosion equipment is a good illustration, Figure A6. 1. In that case, the pretreatment system consists of 1) a biomass receiver where hot water is added, 2) a steam preheater and, 3) a horizontal pretreatment reactor, operating at high pressures and short residence times. Acid addition occurs at the plug feeders transporting the biomass from the pre-steamer to the pretreatment reactor. This equipment is designed for working at 30% total solids loading and 190 °C, with 18 mg/g dry biomass of acid load, discharging the biomass to a flash vessel set at 130 °C (Humbird *et al.*, 2011).

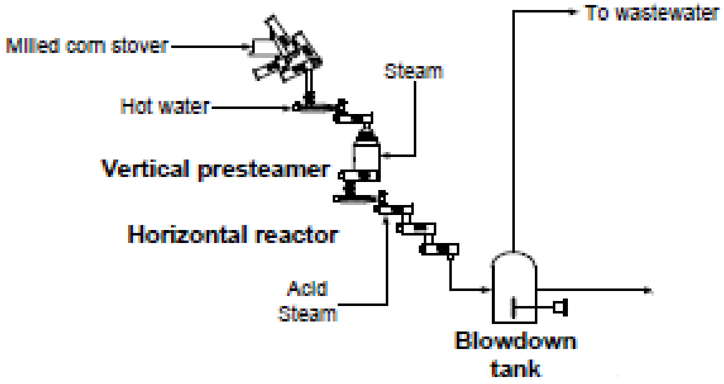


Figure A6. 1. Scheme of the pretreatment set-up for switchgrass, adapted from Humbird et al. (2011).

After reviewing the setting for the pretreatment system in NREL's 2011 report, and considering that the biomass entering the pretreatment step has around 50% moisture content (in comparison to dried switchgrass used as basis for the NREL design), a biomass receiver with water addition is not considered necessary. Thus, the pretreatment equipment considered for this process is composed of: 1) a preheater, in which steam is injected to raise the wet biomass temperature to 100 °C, increasing at the same time the moisture content of the biomass entering the pretreatment reactor, and 2) a horizontal high pressure reactor at 190 °C, into which the biomass and high pressure steam are added. Biomass feed is delivered to the horizontal reactor though plug screw feeders, at the end of which the acid solution is delivered. The biomass is exploded by discharging the slurry into a flash tank receiver to atmospheric pressure. The conveyors utilized for the preparation of the biomass can be arranged to deliver the biomass to the initial plug screw feeder for the preheater.

1.4. Fibres separation

Equipment for dewatering of biomaterial is commonly used in waste water and pulp and paper industries. Typical technologies used in these industries are: 1) belt press, 2) screw press and, 3) rotary press. After reviewing the advantages and disadvantages related to these three options (Table A6. 4), it is clear that rotary press carries more advantages. However, this type of press is typically intended for wastewater secondary sludge, with typical solids loadings much lower than 10% (w/w), which is pumped into the equipment. Considering the pretreated slurry is 30% total solids, rotary press is not an alternative. Thus, screw press is selected as the most suitable equipment for separating the pretreated slurry.

Table A6. 4. Comparison of dewatering equipment, adapted from TOMOE Engineering (2007)

	Belt Press	**Screw Press**	**Rotary Press**
Structure			
Principle:	Filtration	Filtration	Filtration
Driving force:	Gravity	Pressure	Pressure
Separation:	Gravitational filtration	Pressurized filtration	Pressurized filtration
Filter:	Filter cloth	Perforated metal	Perforated metal
Discharge:	Running filter	Screw conveyor	Friction
Advantage	Less noise High soluble solids recovery	Less noise Less rinse water	Energy saving Lighter Less noise Less rinse water
Disadvantage	Need filter change Equipment corrosion Need filter cleansing Many accessories Heavy equipment	Low soluble solids recovery Need cleansing of outer casing	Fewer sales records

Screw presses rely on a large internal screw that pulls incoming material along a horizontal tube. At the end of the tube, a plate impedes the movement of the solids and the generated pressure on the solid plug forces the liquid out of the stream. The screw finally pushes the dewatered solids past the plate onto a collection carrier, Figure A6. 2. Moisture contents achieved with screw press are reported in the range of 65% (Vincent Corporation, 2012) and are thus considered in the model.

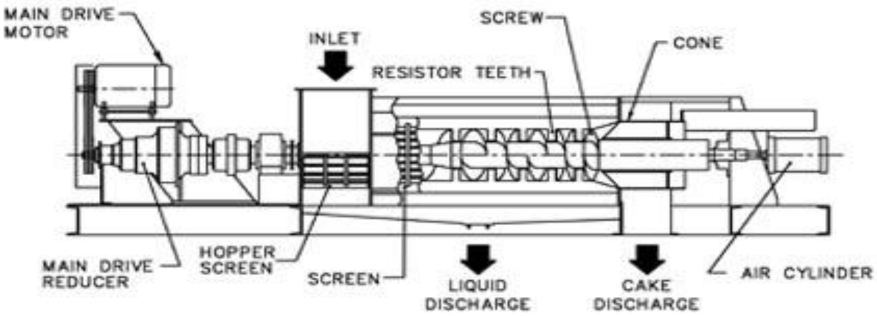


Figure A6. 2. Twin screw press scheme, taken from Vincent Corporation (2012)

1.5. Enzymatic hydrolysis

Several enzyme preparations are in the market, some of these supply the enzymes to their own producers, like Abengoa in partnership with Dyadic (table A6.5).

Table A6. 5. Enzyme producers and second generation chemical producers, adapted from Lane (2012)

Producer	Novozymes	Genencor	Abengoa	DSM	BP	Mascoma
Fiberight	x					
Poet				x		
DCE (Dupont)		x				
Abengoa			x			
BP					x	
Mascoma						x
Petrobras	x					
Inbicon	x	x				
Chemtex	x					
Corfco	x					

From the enzyme producers in Table A6. 5, Genencor and Novozymes A/S are broadly recognized enzyme producers. The DuPont-Genencor venture released the cellulosic enzyme preparation Accellerase TRIO in 2011. On the other hand, Novozymes remains as an enzyme focused company, and in 2012 released their newest cellulosic enzyme preparation, Cellic CTec3. Novozymes Cellic CTec3 is a mixture of cellulases and hemicelluloses. For pretreatments in which hemicellulose hydrolysis is reached in a relatively high level, this enzyme preparation is considered sufficient for digesting remaining hemicellulose.

Optimal conditions are similar for commercial cellulosic enzyme preparations. Novozymes Cellic CTec3 has an optimum temperature at 50-55°C and optimum pH in the range of 4.75 and 5.25. Considering that NREL studies on several commercial enzymes preparations show optimum temperature as 48 °C, 50 °C is selected for hydrolysis temperature (Humbird *et al.*, 2011).

Enzymes loadings for a given conversion depend greatly on the type of substrate and pretreatment it undergoes. Suggested enzyme loadings by Novozymes for Cellic CTec3 range from 1 to 6% w/w (enzyme preparation/cellulose). According to Novozymes internal studies on acid pretreated corn stover (solids loadings of 18% w/w), loadings of 4% Cellic CTec3 can be used to reach 90% cellulose conversion (Novozymes A/S, 2012).

Enzymatic hydrolysis equipment type depends greatly on the material to hydrolyse. For high solids mixtures (> 15 %TS), there are several equipment examples at industrial scale, particularly in the anaerobic digestion of municipal solid waste (MSW). After a review of digestion systems for solid residues the following type of configurations were found

- One stage leach-bed systems: The solid waste is simply contained in a closed and anaerobic environment while it undergoes the complete digestion process. A leachate is collected at the bottom and can be recirculated back onto the solids.

- Plug-flow reactors: In these cases, part of the treated waste is mixed with the incoming fresh waste. The recirculation and mixing is used for inoculation and to prevent possible overloading. Examples are the Dranco process, where the solids slowly moves down due to gravity while it is simultaneously degraded. In the Kompogas process the flow occurs horizontally, and is aided by slow-rotating impellers (Vandevivere et al., 2002).

- Two-stage systems: This type of systems typically consist of a leach-bed reactor followed by a methanogenic reactor for biogas production. Large scale examples of these systems are the Bio Vækst and Hera Vekst plants in Denmark and Norway. These plants are based on the

AIKAN system in which well mixed MSW is loaded into anaerobic modules where liquid from the methanogenic reactor is spread back onto the solids (Solum-Gruppen).

The enzymatic hydrolysis for the pretreated biomass residues, no inoculum is necessary and thus there is no need to recirculate the solids. Thus, parallel vertical plug-flow reactors are a suitable simple straightforward approach. The reactors can be vertical towers in which the biomass flows downwards as it is continuously hydrolysed, similar to those described in 2002 NREL report (Aden *et al.*, 2002)

1.6. Heat and Power Generation

Different technologies are available for generation of heat and power, each with its own advantages and disadvantages, see Table A6. 6. After reviewing the available technology, it becomes clear that steam turbine systems are the best option. First, steam turbines allows the feed of solid and gas fuels, since the energy is transferred separately, first from the fuel to steam and then to the turbine. Secondly, steam turbines can have multiple extraction points, allowing for recovery of steam at different conditions (low and/or medium pressure steam), which is best suited to supply energy to the adjacent mill and the design process.

Table A6. 6. Comparison of prime mover technologies applicable to biomass, adapted from U.S. Environmental Protection Agency (2008).

Characteristic	Steam Turbine	Gas Turbine	Micro-turbine	Reciprocating IC Engine	Fuel Cell	Stirling Engine
Size	50 kW – 250 MW	500 kW – 40 MW	30 kW – 250 kW	<5 MW	<1 MW	< 200kW
Fuels	Biomass/biogas	Biogas	Biogas	Biogas	Biogas	Biomass/biogas
Fuel preparation	None	Filter	Filter	Filter	Sulphur, CO issues	None
Sensitive to fuel moisture	N/A	Yes	Yes	Yes	Yes	No
Electric efficiency *	5 – 30 %	22 – 36%	22 – 30%	22 – 45%	30 – 63%	5 – 45%

Operating issues	High reliability, slow start-up, long life, readily available.	High reliability, no cooling required, requires gas compressor	Fast start-up, requires fuel gas compressor	Fast start-up, must be cooled when CHP heat is not used.	Low durability, low noise	Low noise
Field experience	Extensive	Extensive	Extensive	Extensive	Some	Limited
Installed cost (USD/kW)	350 – 750**	700 – 2 000	1 100 – 2 000	800 - 1500	3 000 – 5 000	1 000 – 10 000
Operation and maintenance cost (USDc/kWh)	< 0.4	0.6 – 1.1	0.8 – 2.0	0.8 – 2.5	1 – 4	~ 1

*From high heating value (HHV) of the fuel, considered as the ratio of the incoming energy to that available as electricity.

**Without boiler.

APPENDIX 7. Economic Evaluation Data

1. Total Capital Investment

1.1. Individual Equipment Cost

1.1.1. *Area 100 - Pre-processing*

Pre-processing equipment costing is based on the required processing throughput for each unit, defined in Chapter 7 – Process and Equipment Design. The equipment in this processing area is basically for mechanical conversion, including shredders, debarkers and milling tandems.

Costing of the milling tandem is calculated separately for the crushers and mill composing each tandem. The processing capacity of each tandem is 53 tonne/h. Reported price fir crushers with the required capacity is 50 000 USD (Peters *et al.*, 2003). Prices for milling tandems are estimated on the same source, which in turn result in 200 000 USD per tandem, considering the four mill sets in each tandem. The cost difference amongst both cases depends thus on the number of tandems required for the processing throughput.

The price of shredders and debarkers were not found on literature and thus the cost quoting is referenced to private vendors found on public trading sites. Shredders are found with maximum capacities in the range of 30 tonnes/h, and thus several units are considered for covering the processing requirements. The pricing for this type of equipment is listed in the range of 18 000 USD (Alibaba.com, 2012). Debarking equipment cost with similar capacity is reported in the range of 20 000 USD. Vibrating screens to be coupled after the milling tandems are also quoted from public vendors. Equipment cost for processing 180 tonne/h is in the range of 10 000 USD (Alibaba.com, 2012).

1.1.2. *Area 200 – Biomass Conversion*

The most expensive single equipment is the pretreatment system, which consists of a pre-steamer E201, a screw feeder M201, a pretreatment reactor R201, and a flash tank separator S201. Similar equipment is available through vendors like Andritz. For this project, the base cost for the pretreatment system is taken from NREL's 2011 report, quoted from Andritz at 19 812 448 USD for 87 tonne/h capacity (Humbird *et al.*, 2011). The scaling index is taken as 0.6 from the same source.

The conditioning tank cost is based on required volume for the equipment and is based on stainless steel agitated tank prices given by Peters *et al.* (2003), estimated as 80 000 USD for 65m^3. For the ten-mill case, the cost is scaled with an index of 0.7, as given in NREL's 2011 report for tanks and reactors (Humbird *et al.*, 2011).

Costing of the hydrolysis system is based on the system cost reported in NREL's 2002 report, considering the similarities of the equipment. The base cost is 493 391 USD for hydrolysis tanks of 3 596 m^3 (Aden *et al.*, 2002). The scaling index used for this equipment is 0.7 as used in NREL's 2011 report (Humbird *et al.*, 2011).

Screw press costing is based on public vendors quotes for dewatering screw presses. The maximum capacities found are in the range of 140 tonne/h at 40 000 USD, thus several units are considered to cover the processing requirements.

1.1.3. *Area 300 – Waste Water Treatment*

The costing from the waste water treatment anaerobic rector is based on a concrete anaerobic reactor constructed on-site. The base price is taken as 27 000 000 USD for processing 209 tonne/h of waste waters (Humbird *et al.*, 2011). The concrete aerobic basin is quoted as 10% of the anaerobic reactor cost, (Humbird *et al.*, 2011), and the scale index is taken as 0.6 for both equipment, as reported in the same source.

The clarifier for the treated water is quoted based on the transversal area of the equipment, considering typical overflow rates and the designed outflow (Perry *et al.*, 1997). The quoted cost for equipment with the specified diameter for the three-mill case including installation cost, is listed as 2000 USD/m^2, whereas the cost of larger equipment for the ten-mill case is listed as 1500 USD/m^2.

1.1.4. *Area 400 – Heat and Power Generation*

The main equipment of the heat and power generation area is the furnace/boiler and the steam extraction turbine. This equipment is highly costly, and is quoted based on the electrical power output for the turbine and the mass flow for the boiler. The base price for the system is taken as reported in NREL's 2011 report, as it is a system that already includes the cost of preheater and secondary equipment needed for both biogas and solid fuels (Humbird *et al.*, 2011). The turbine equipment is quoted as 9 381 215 USD for 41 MWe, whereas the boiler is quoted as 28 550 000 USD for 240 tonne/h of feed water. The cost of both types of equipment is scaled using an index of 0.6 as reported in the same source.

Heat exchangers were quoted based on transfer area required. The equipment cost were quoted first to the three-mill case from Peters *et al.* (2003), and then scaled up to the ten-mill case using the scaling index of 0.7 , as given in NREL's 2011 report for tanks and reactors (Humbird *et al.*, 2011).

1.2. Scaled Present Time Equipment Costs

Each equipment cost mentioned in the previous section, is referenced from different sources and in different historical times. In order to homogenize past currency units, the equipment costs from different years are converted into 2011 U.S. dollars (USD), following equation A7.1. Inflation values from the United States are taken as reported by U.S. Bureau of Labor Statistics (2012), listed in table A7.1. The equipment obtained from vendors in the current year is left in 2012 USD as no average inflation index is available yet.

$$\mathrm{Cost}_{2011} := \mathrm{Cost}_{\mathrm{Base}} \cdot \left(\frac{\mathrm{Index}_{2011}}{\mathrm{Index}_{\mathrm{Base}}} \right)^{\blacksquare} \qquad \text{(equation A7.1)}$$

Table A7. 1. U.S. Inflation Indices

Year	Inflation Index
1997	160.50
1998	163.00
1999	166.60
2000	172.20
2001	177.10
2002	179.90
2003	184.00
2004	188.90
2005	195.30
2006	201.30
2007	207.34
2008	215.30
2009	214.54
2010	218.06
2011	224.94

The pricing for equipment is based on capacity. In some cases the reference cost capacity can be scaled to match the processing requirement. The scaling of equipment cost is estimated following equation A7.2, using the scaling indices as mentioned for each equipment in the previous section. In some cases, an equipment unit has a maximum capacity, for which scaling is not considered but rather several units are necessary.

$$\mathrm{Cost}_{\mathrm{Scale2}} := \mathrm{Cost}_{\mathrm{Scale1}} \cdot \left(\frac{\mathrm{Capacity}_2}{\mathrm{Capacity}_1}\right)^{i_{\mathrm{scale}}} \qquad \text{(equation A7.2)}$$

Where,
i_{scale} = Scale index for each equipment, see table A7.2.

Table A7. 2. Three-mill Case Equipment Purchase Cost

Equipment		Required Capacity		Base Cost			Scale Index *	Scaled Base Cost	Scaled Actual Cost
Code	Name	-	Unit	USD x 10^3	Cap.	Year		USD x 10^3	USD x 10^3
A101	OPF Shredder	110	tonne/h	110	tonne/h	2012	-	18	18
A102	OPT Debarker	90	tonne/h	90	tonne/h	2012	-	20	20
A103	OPT Shredder	78	tonne/h	78	tonne/h	2012	-	18	18
A104a	Premill Crusher	53	tonne/h	53	tonne/h	1991	-	50	70
A104	Mill tandems	53	tonne/h	53	tonne/h	1991	-	200	280
A105	EFB Shredder	41	tonne/h	41	tonne/h	2012	-	18	18
S101	Vibrating Screen	178	tonne/h	178	tonne/h	2002	-	10	13
E201/ R201/ S201	Pretreatment System	75	dry tonne/h	75	dry tonne/h	2009	0.60	18 124	19 003
E202	Heat Exchanger	680	m2	680	m2	2002	-	102	128
E203	Vent Condenser	1494	m2	1494	m2	2002	0.70	187	234
E204	Heat Exchanger	1881	m2	1881	m2	2002	0.70	208	260
E205	Heat Exchanger	35	m2	35	m2	2002	0.70	14	17
E206	Heat Exchanger	680	m2	680	m2	2002	0.70	102	128
R202	Conditioning Tank	65	m3	65	m3	2002	-	80	100
R203 / M203	Hydrolysis System	3500	m3	3500	m3	1998	0.70	484	668
S202	Screw Press	253	tonne/h	253	tonne/h	2012	-	40	40
S203	Rotary Drum Filter	180	m2	180	m2	1997	0.48	262	368
S204	Screw Press	380	tonne/h	380	tonne/h	2012	-	40	40
T201	Acid Feeding Tank	9	m3	9	m3	2002	-	30	38
T202	Enzyme Feeding Tank	13	m3	13	m3	2002	-	35	44
T203	Lime Feeding Tank	1	m3	1	m3	2002	-	10	13
T204	Sugar Tank	58	m3	58	m3	2002	-	70	88
E301	Heat Exchanger	33	m2	33	m2	2002	-	8	10
R301	Anaerobic Reactor	90	tonne/h	90	tonne/h	2010	0.60	11 165	11 517
R302	Aerobic Basin	90	tonne/h	90	tonne/h	2010	0.60	1 116	1 152

S301	Clarifier	78	m2	78	m2	1997	0.00	141	197
TG401	Turbine System	19	MWe	19	MWe	2010	0.60	5 960	6 148
F401/ Boiler	Boiler	134	tonne/h	134	tonne/h	2010	0.60	20 192	20 829

* Scale indices are taken from NREL 2011 report (Humbird *et al.*, 2011), except the index for S203, which is taken from Perry *et al.* (1997).

Table A7. 3. Ten-mill Case Equipment Cost

Equipment		Req. Capacity		Base Cost			Scale index *	Scaled Based Cost	Scaled Actual Cost
Code	Name	-	Unit	10^3 x USD	Cap.	Year	-	10^3 x USD	10^3 x USD
A101	OPF Shredder	367	tonne/h	18	30	2012	-	18	18
A102	OPT Debarker	300	tonne/h	20	30	2012	-	20	20
A103	OPT Shredder	261	tonne/h	18	30	2012	-	18	18
A104a	Premill Crusher	53	tonne/h	50	53	1991	0.00	50	70
A104	Mill tandems	53	tonne/h	200	53	1991	0.00	200	280
A105	EFB Shredder	138	tonne/h	18	30	2012	-	18	18
S101	Vibrating Screen	596	tonne/h	10	178	2002	0.86	28	35
E201/ R201/ S201	Pretreatment System	250	dry tonne/h	19 812	87	2009	0.60	37 324	39 134
E202	Heat Exchanger	1212	m2	102	680	2002	0.70	153	191
E203	Vent Condenser	2643	m2	119	783	2002	0.70	279	349
E204	Heat Exchanger	6280	m2	102	680	2002	0.70	483	605
E205	Heat Exchanger	118	m2	119	783	2002	0.70	32	39
E206	Heat Exchanger	2297	m2	102	680	2002	0.70	239	299
R202	Conditioning Tank	225	m3	80	65	2002	0.70	191	239
R203/ M203	Hydrolysis System	4395	m3	493	3596	1998	0.70	568	784
S202	Screw Press	845	tonne/h	40	140	2012	-	40	40
S203	Rotary Drum Filter	180	m2	63	9	1997	0.48	262	368
S204	Screw Press	1274	tonne/h	40	140	2012	-	40	40
T201	Acid Feeding Tank	30	m3	30	9	2002	0.70	73	91
T202	Enzyme Feeding Tank	45	m3	35	13	2002	0.70	86	107
T203	Lime Feeding Tank	4	m3	10	1	2002	0.70	22	27
T204	Sugar Tank	180	m3	70	55	2002	0.70	161	201
E301	Heat Exchanger	29	m2	8	33	2002	0.70	7	9
R301	Anaerobic Reactor	209	tonne/h	27 000	393	2010	0.60	18 485	19 068
R302	Aerobic Basin	209	tonne/h	2 700	393	2010	0.60	1 848	1 907
S301	Clarifier	180	m2	243	180	1997	0.00	243	340
TG401	Turbogenerator System	61	MWe	9 381	41	2010	0.60	11 850	12 225
F401/ Boiler	Boiler	428	tonne/h	28 550	239	2010	0.60	40 530	41 809

* Scale indices are taken from NREL 2011 report (Humbird *et al.*, 2011), except for S203, which is taken from Perry *et al.* (1997) and heat exchanging equipment from (Peters *et al.*, 2003).

1.3. Total Bare Module Cost

After quoting the equipment necessary for the process, the installed equipment cost was calculated taking into account the number of units required for covering the process throughput, and an installation factor. It was decided to use an installation factor that accounts for both installation of the main equipment, as well as for instrumentation and control, and piping and cost. For few equipment units, the installation factor is equal to one; in those cases the quoted price of the equipment already includes these costs. The factors are referenced to NREL 2011 report, and are based on separate installation instrumentation factors or on specific quotations from vendors (Peters *et al.*, 2003; Humbird *et al.*, 2011). Although in NREL study does not include crushing or milling operations, the factor used for solid handling equipment in the mentioned report is considered in good accordance with typical factors assigned to these types of equipment.

Table A7. 4. Three-mill Case Installed Equipment Cost

Equipment		Scaled Actual Cost	Units	Total Purchased Cost	Inst. Factor	Total Bare Module Cost	
Code	Name	USD	No.	USD	-	USD x 10^3	RM x 10^3
A101	OPF Shredder	18 000	4	72 000	1.7	122	378
A102	OPT Debarker	20 000	3	60 000	1.7	102	315
A103	OPT Shredder	18 000	3	54 000	1.7	92	283
A104a	Premill Crusher	70 074	4	280 298	1.7	477	1 471
A104	Mill tandem	280 298	4	1 121 191	1.7	1 906	5 883
A105	EFB Shredder	18 000	2	36 000	1.7	61	189
S101	Vibrating Screen	12 504	1	12 504	1.7	21	66
E201/ R201/ S201	Pretreatment System	19 003 163	1	19 003 163	1.5	28 505	87 978
E202	Heat Exchanger	127 536	1	127 536	2.2	281	866
E203	Vent Condenser	233 892	1	233 892	2.2	515	1 588
E204	Heat Exchanger	259 986	1	259 986	2.2	572	1 765
E205	Heat Exchanger	17 006	1	17 006	2.2	37	115
E206	Heat Exchanger	127 567	1	127 567	2.2	281	866
R202	Conditioning Tank	100 028	1	100 028	1.8	180	556
R203/M 203	Hydrolysis System	668 101	6	4 008 607	2	8 017	24 744
S202	Screw Press	40 000	2	80 000	1.7	136	420
S203	Rotary Drum Filter	367 834	3	1 103 503	1.7	1 876	5 790
S204	Screw Press	40 000	3	120 000	1.7	204	630
T201	Acid Feeding Tank	37 511	1	37 511	1.8	68	208
T202	Enzyme Feeding Tank	43 762	1	43 762	1.8	79	243
T203	Lime Feeding Tank	12 504	1	12 504	1.8	23	69

T204	Sugar Tank	87 525	1	87 525	1.8	158	486
E301	Heat Exchanger	10 003	1	10 003	2.2	22	68
R301	Anaerobic Reactor	11 517 252	1	11 517 252	1	11 517	35 547
R302	Aerobic Basin	1 151 725	1	1 151 725	1	1 152	3 555
S301	Clarifier	197 368	1	197 368	1	197	609
TG401	Turbine System	6 147 739	1	6 147 739	1.8	11 066	34 154
F401/ Boiler	Boiler	20 829 054	1	20 829 054	1.8	37 492	115 717

Table A7. 5. Ten-mill Case Installed Equipment Cost

Equipment		Scaled Actual Cost	Units	Total Purchased Cost	Inst. Factor	Total Bare Module Cost	
Code	Name	USD	No.	USD	-	USD x 10^3	RM x 10^3
A101	OPF Shredder	18 000	13	234 000	1.7	398	1 228
A102	OPT Debarker	20 000	10	200 000	1.7	340	1 049
A103	OPT Shredder	18 000	9	162 000	1.7	275	850
A104a	Premill Crusher (12 tandems)	70 074	12	840 893	1.7	1 430	4 412
A104	Mill tandems (12 tandems)	280 298	12	3 363 574	1.7	5 718	17 648
A105	EFB Shredder	18 000	5	90 000	1.7	153	472
S101	Vibrating Screen	35 350	1	35 350	1.7	60	185
E201/ R201/ S201	Pretreatment System	39 133 902	1	39 133 902	1.5	58 701	181 175
E202	Heat Exchanger	191 141	1	191 141	2.2	421	1 298
E203	Vent Condenser	348 689	1	348 689	2.2	767	2 368
E204	Heat Exchanger	604 547	1	604 547	2.2	1 330	4 105
E205	Heat Exchanger	39 449	1	39 449	2.2	87	268
E206	Heat Exchanger	298 980	1	298 980	2.2	658	2 030
R202	Conditioning Tank	238 568	1	238 568	1.8	429	1 325
R203/ M203	Hydrolysis System	783 548	16	12 536 774	2	25 074	77 387
S202	Screw Press	40 000	6	240 000	1.7	408	1 259
S203	Rotary Drum Filter	367 834	9	3 310 510	1.7	5 628	17 370
S204	Screw Press	40 000	10	400 000	1.7	680	2 099
T201	Acid Feeding Tank	90 687	1	90 687	1.8	163	504
T202	Enzyme Feeding Tank	107 279	1	107 279	1.8	193	596
T203	Lime Feeding Tank	27 461	1	27 461	1.8	49	153
T204	Sugar Tank	200 710	1	200 710	1.8	361	1 115
E301	Heat Exchanger	9 128	1	9 128	2.2	20	62
R301	Anaerobic Reactor	19 068 333	1	19 068 333	1	19 068	58 853
R302	Aerobic Basin	1 906 833	1	1 906 833	1	1 907	5 885
S301	Clarifier	340 290	1	340 290	1	340	1 050
TG401	Turbine System	12 224 534	1	12 224 534	1.8	22 004	67 914

F401/ Boiler	Boiler	41 809 303	1	41 809 303	1.8	75 257	232 274

2. Total Production Cost

Total production cost (TPC) are calculated as a sum of direct production cost (DPC), overhead and general expenses. DPC are composed of raw material cost, labour, maintenance and a capital charge. Raw material costs are distinguished between biomass and other materials, as biomass changes prices according to the case. Labour is calculated assuming a wage of 1800 RM/month and 16 operators in the three mill case. Labour for the ten-mill case is scaled by changing the number of operators required, assuming a scaling index of 0.25 (Peters *et al.*, 2003). The contributions to the total labour, maintenance and other production cost are listed in tables A7.7 and A7.8, (Seider *et al.*, 2010).

Table A7. 6. Three-mill Case - Cost Contributions in Total Production Cost

Cost Contribution		RM/y	RM/tonne
Capital Charge	C_{cap}	63 422 970	178
Biomass	C_{BM}	174 838 960	490
Raw Materials	C_{RM}	13 866 210	39
Waste Disposal	C_{waste}	19 200	0
Labour Wages		2 808 000	8
Supervision Salaries		421 200	1
Supplies and Services		168 480	0
Technical Assistance to Manufacturing		97 200	0
Control Laboratory		105 300	0
Total Labour Cost	C_{lab}	3 600 180	10
M. Wages and Benefits		20 517 163	57
M. Salaries and Benefits		5 129 291	14
M. Materials and services		20 517 163	57
Total Maintenance Cost	C_{maint}	46 163 617	129
Mechanical Department Services		693 016	2
Employee Relations Department		1 703 664	5
Business Services		2 136 798	6
Overhead	C_{overh}	4 533 478	13
Insurance and Taxes	C_{tax}	9 118 739	26
Selling/transfer expense		19 037 108	53
Direct research		30 459 372	85
Allocated research		3 172 851	9
Administrative expense		12 691 405	36
Management incentive expense		7 932 128	22
General Expenses	C_{gral}	73 292 864	205
TOTAL PRODUCTION COST	**TPC**	**388 856 217**	**1089**

Table A7. 7. Ten-mill Case - Cost Contributions in Total Production Cost

Cost Contribution		RM/y	RM/tonne
Capital Charge	C_{cap}	136 101 599	114
Biomass	C_{BM}	787 482 550	659
Raw Materials	C_{RM}	48 551 806	41
Waste Disposal	C_{waste}	67 200	0
Labour Wages		3 861 000	3
Supervision Salaries		579 150	0
Suuplies and Services		231 660	0
Technical Assistance to Manufacturing		97 200	0
Control Laboratory		105 300	0
Total Labour Cost	C_{lab}	4 874 310	4
M. Wages and Benefits		44 028 508	37
M. Salaries and Benefits		11 007 127	9
M. Materials and services		44 028 508	37
M Overhead		2 201 425	2
Total Maintenance Cost	C_{maint}	101 265 568	85
General Plant Overhead		4 222 781	4
Mechanical Department Services		1 427 419	1
Employee Relations Department		3 509 071	3
Business Services		4 401 208	4
Overhead	C_{overh}	13 560 479	11
Insurance and Taxes	C_{tax}	19 568 226	16
Selling/transfer expense		63 658 752	53
Direct research		101 854 002	85
Allocated research		10 609 792	9
Administrative expense		42 439 168	36
Management incentive expense		26 524 480	22
General Expenses	$C_{gra}l$	245 086 193	205
TOTAL PRODUCTION COST	**TPC**	**1 356 557 930**	**1136**

3.Cash Flows Analysis

Incoming cash flows after the first operational year of the plant are calculated following equation A7.3. Outgoing flows are the calculated TCI in the previous section. It is assumed that TC excluding working capital is equally spent in the two construction years, while the working capital is only considered at the end of the second year. All the incoming and outgoing cash flows are converted into present value taking into consideration the current interest rate in Malaysia, equation A7.4., and are presented in tables A7.8 and A7.9.

$$CF := (1 - t)\cdot\left[\text{Sales}_{\text{revenue}} - \left(TPC - C_{cap}\right)\right] + t\cdot\text{depreciation} \qquad \text{(equation A7.3)}$$

Where:
t = taxes on corporate income in Malaysia, 25%

$$PV := CF\cdot(1 + i)^{-n} \qquad \text{(equation A7.4)}$$

Where:
PV = Present Value
CF = Cash Flow
n = year after present time

Table A7. 8. Cash Flows from Three-mill Case

Project time (Year)	CF (RM)	CF Present Value (RM)	CF Net Present Value (RM)
1	-227 968 480	-214 054 911	-214 054 911
2	-308 213 385	-271 739 192	-485 794 103
3	280 056 315	231 844 366	-253 949 737
4	280 056 315	217 694 240	-36 255 497
5	280 056 315	204 407 737	168 152 241
6	280 056 315	191 932 148	360 084 388
7	280 056 315	180 217 979	540 302 368
8	280 056 315	169 218 760	709 521 127
9	280 056 315	158 890 854	868 411 982
10	280 056 315	149 193 290	1 017 605 272
11	280 056 315	140 087 597	1 157 692 869
12	280 056 315	131 537 649	**1 289 230 518**

Table A7. 9. Cash Flows from Ten-mill Case

Project time (Year)	CF (RM)	CF Present Value (RM)	CF Net Present Value (RM)
1	-489 205 643	-459 348 021	-459 348 021
2	-661 406 029	-583 134 765	-1 042 482 787
3	820 510 642	679 258 990	-363 223 797
4	820 510 642	637 801 869	274 578 072
5	820 510 642	598 874 994	873 453 066
6	820 510 642	562 323 938	1 435 777 004
7	820 510 642	528 003 698	1 963 780 702
8	820 510 642	495 778 120	2 459 558 822
9	820 510 642	465 519 361	2 925 078 184
10	820 510 642	437 107 382	3 362 185 565
11	820 510 642	410 429 466	**3 772 615 032**

APPENDIX 8. Life Cycle Inventories

1. MICCI Palm Oil inventories

1.1. Plantation Inventory

The interventions related to the plantation of oil palm are listed in table A8.1. These interventions are based on the LCI published by Schmidt (2007) on Malaysian and Indonesian palm oil production. The LCI data used for each intervention is in most cases the same as those used by the same author, although in some cases other databases had to be used due to differences in database access. The only structural difference considered in the LCI of the plantation for feeding the MICCI cluster is that OPT and OPF are considered co-products. No other differences were considered as (1) the carbon emissions reported for a conventional mill are based on soil changes and land transformation, without considering biotic emissions, (2) nutrient inputs from biomass and fertilizers are accounted together already in the conventional mill inventory, and the same amounts of nutrients are necessary for the same plantation area regardless of the intended use of cropped biomass.

Table A8. 1 Plantation inventory for MICCI cluster

Inventory			Applied LCI Data
Outputs			
FFB	tonne/ha/y	18.87	
OPT		9.44	
OPF (1/3)		11.53	
Material use			
N-Fertiliser Ammonium Sulphate	kg/ha/y	76.00	Ammonium Sulphate, as N, at regional storehouse, RER U
N-fertiliser Urea	kg/ha/y	28.00	Urea, as N, at regional storehouse
P-Fertiliser	kg/ha/y	70.00	Fertiliser
K-Fertiliser	kg/ha/y	204.00	Potassium chloride, as K_2O, at regional storehouse
Herbicide	kg/ha/y	2.40	Glyphosate, at regional storehouse
Insecticide	kg/ha/y	0.32	Pesticide unspecified, at regional storehouse
Energy use			
Traction, burned diesel	MJ/ha/y	2118.00	Diesel
Electricity for overhead	MJ/ha/y	0.05	
Capital goods			
Agricultural buildings	m2/ha/y	0.04	Shed
Administration building	m3/ha/y	0.00	Building, multi-storey
Machinery, tractor	kg/ha/y	4.40	Tractor, production
Machinery, tillage	kg/ha/y	0.40	Agricultural machinery, tillage, production
Machinery, general	kg/ha/y	2.10	Agricultural machinery, general, production
Transport			

Transport, transoceanic	tkm/ha/y	11520.00	Transport, transoceanic tanker
Transport, lorry	tkm/ha/y	1152.00	Transport, lorry >28t, fleet average
Emissions to air			
Carbon dioxide	kg/ha/y	1500.00	
Ammonia	kg/ha/y	18.30	
Dinitrogen monoxide	kg/ha/y	10.10	
Nitric oxide	kg/ha/y	3.20	
Glyphosate	kg/ha/y	0.80	
Cypermethrin	kg/ha/y	0.10	
Emissions to water			
Nitrate	kg/ha/y	353.00	
Phosphorus	kg/ha/y	1.60	
Glyphosate	kg/ha/y	0.80	
Cypermethrin	kg/ha/y	0.10	
Emissions to soil			
Glyphosate	kg/ha/y	0.80	
Cypermethrin	kg/ha/y	0.10	
Arsenic	g/ha/y	1.00	
Cadmium	g/ha/y	1.90	
Chromium	g/ha/y	44.00	
Cobalt	g/ha/y	0.23	
Copper	g/ha/y	12.00	
Mercury	g/ha/y	0.03	
Molybdenum	g/ha/y	0.11	
Nickel	g/ha/y	5.10	
Lead	g/ha/y	2.00	
Selenium	g/ha/y	0.42	
Zink	g/ha/y	69.00	

1.2. Palm Kernel Milling Inventory

The interventions related to palm kernel oil extraction are taken from the inventory published by Schmidt (2007), table A8.2. These interventions are considered equal to those related to any conventional palm oil mill and thus there are no changes in the inventory.

Table A8. 2 Palm kernel oil milling inventory for MICCI cluster.

Inventory			Applied LCI Data
Outputs			
PKO	kg/tonneCPKO	1000	
PKC	kg/tonneCPKO	1161	
Material Use			
Kernel	kg/tonneCPKO	2228	
Water	kg/tonneCPKO	400	Water, river
Energy use			
Electricity	MJ/tonneCPKO	755	Hard coal, burned in power plant *
Emissions			
Nitrate to water	g/tonneCPKO	3	
Capital goods			
Building halls	m2/tonneCPKO	8.60E-04	Building, hall, steel construction
Building, multi story	m3/tonneCPKO	1.60E-03	Building, multi-storey
Machinery	kg/tonneCPKO	0.19	Facilities, chemical production
Transport			
Transport, lorry	tkm/tonneCPKO	222	Transport, lorry 20-28t, fleet average

*Multiplied by the efficiency of burning coal, considering Malaysian power plants, as used by Schmidt, 2007.

1.3. Enzyme Production Inventory

Enzyme production inventory is taken as reported in the U.S. LCI Database (National Renewable Energy Laboratory, 2012). The interventions are related to the production of Novozymes Celluclast with update data from 2007 and 2010. Transportation of the enzyme to the MICCI facilities is considered negligible given the small quantities in comparison to transport requirements of fertilizers and biomass.

Table A8. 3 Enzyme production inventory for MICCI cluster.

Inventory			Applied LCI Data
Products			
Enzyme, cellulase	kg	1	
Material use			
Water	m3	0.0209	
Land	m2	0.723	Transformation from pasture
Land	m2	0.723	Transformation to industrial area
Energy from coal	MJ	52.1	
Emissions to air			
Carbon dioxide, fossil	kg/kg enzyme	4.09	
Ethene	kg/kg enzyme	0.002	
Sulgur dioxide	kg/kg enzyme	0.0153	
Emissions to soil			
Phosphate	kg/kg enzyme	0.00937	

1.4. MICCI cluster inventory

The inventory including interventions related to palm oil milling, including the proposed process for production of fermentable sugars is presented in table A8.4.

Table A8. 4 Inventory for MICCI cluster.

Intervention			Applied LCI Data
Outputs			
CPO	kg/tonneFFB	199.8	
Kernel	kg/tonneFFB	53.2	
Material Use			
FFB	kg/tonneFFB	1000	
Water	tonne/tonneFFB	3.06	Water, river
OPT	tonne/tonneFFB	0.50	
OPF (1/3)	tonne/tonneFFB	0.61	
Enzymes	kg/tonneFFB	0.13	From U.S. Life Cycle Inventories Database, 2012.
Lime	kg/tonneFFB	6.72	Lime, hydrated, packed, at plant/CH U - ecoinvent
Acid	kg/tonneFFB	9.44	Sulphuric acid, liquid, at plant/RER U - ecoinvent
Displaced products			
Bark	tonne/tonneFFB	0.07	Bark, at rough green lumber sawmill, softwood, US PNW/kg/US
Sand (ash stream)	tonne/tonneFFB	0.03	Sand, at mine/CH U - ecoinvent
Hard coal (shell stream)	kg/tonneFFB	55	Coal - LCA Food DK
FS Product	tonne/tonneFFB	0.33	Sugar, from sugar beet, at sugar refinery/CH U - ecoinvent
Biogas	m3/tonneFFB	0.02	Biogas, from slurry, at agricultural co-fermentation, covered/CH U - Ecoinvent
Energy use			
Steam	tonne/tonneFFB	0.28	On-site steam average R - ecoinvent
Electricity	MJ/tonneFFB	31.80	Hard coal, burned in power plant/NORDEL U - ecoinvent
Diesel, vehicles	MJ/tonneFFB	7.6	Diesel, burned in building machine GLO U - ecoinvent
Waste Treatment			
Gypsum	kg/tonneFFB	12.83	Disposal, gypsum, 19.4% water, to inert material landfill/CH U
Capital Goods			
Building halls	m2/tonneFFB	0.0046	Building, hall, steel construction/CH/I U - ecoinvent
Building, multi-story	m3/tonneFFB	0.0085	Building, multi-storey, RER I U - ecoinvent
Machinery	kg/tonneFFB	2	Facilities, chemical production/RER/I U - ecoinvent
Transport			
Transport diesel	tkm/tonneFFB	0.036	Transport, lorry 20-28t, fleet average/CH U - ecoinvent
Transport	tkm/tonneFFB	0.00	Transport, lorry 20-28t, fleet average/CH U - ecoinvent
Emissions to air			

NOX	g/tonneFFB	9.42E-01	
SO2	g/tonneFFB	8.31E-03	
Ammonia	g/tonneFFB	1.18E-02	
Arsenic	g/tonneFFB	6.59E-06	
Cadmium	g/tonneFFB	4.62E-06	
Chlorine	g/tonneFFB	1.18E-03	
Chromium	g/tonneFFB	2.61E-05	
Cromium VI	g/tonneFFB	2.64E-07	
Copper	g/tonneFFB	1.45E-04	
Dinitrogen monoxide	g/tonneFFB	2.01E-02	
Lead	g/tonneFFB	1.65E-04	
Manganese	g/tonneFFB	1.12E-03	
Mercury	g/tonneFFB	1.98E-06	
Nickel	g/tonneFFB	3.95E-05	
Phenol, pentachloro-	g/tonneFFB	5.34E-08	
Phosphorus	g/tonneFFB	1.98E-03	
Zinc	g/tonneFFB	1.98E-03	
CO	g/tonneFFB	2.32E+00	
Acetaldehyde	g/tonneFFB	4.02E-04	
Benzene	g/tonneFFB	6.00E-03	
Benzene, ethyl	g/tonneFFB	1.98E-04	
Benzo(a)pyrene	g/tonneFFB	3.30E-06	
Formaldehyde	g/tonneFFB	8.58E-04	
Hydrocarbons, aliph. alkanes	g/tonneFFB	6.00E-03	
Hydrocarbons, alphatic	g/tonneFFB	2.08E-02	
m-xylene	g/tonneFFB	7.89E-04	
Methane	g/tonneFFB	4.62E-03	
PAH	g/tonneFFB	7.34E-05	
Toluene	g/tonneFFB	1.98E-03	

2. Life Cycle Assessment

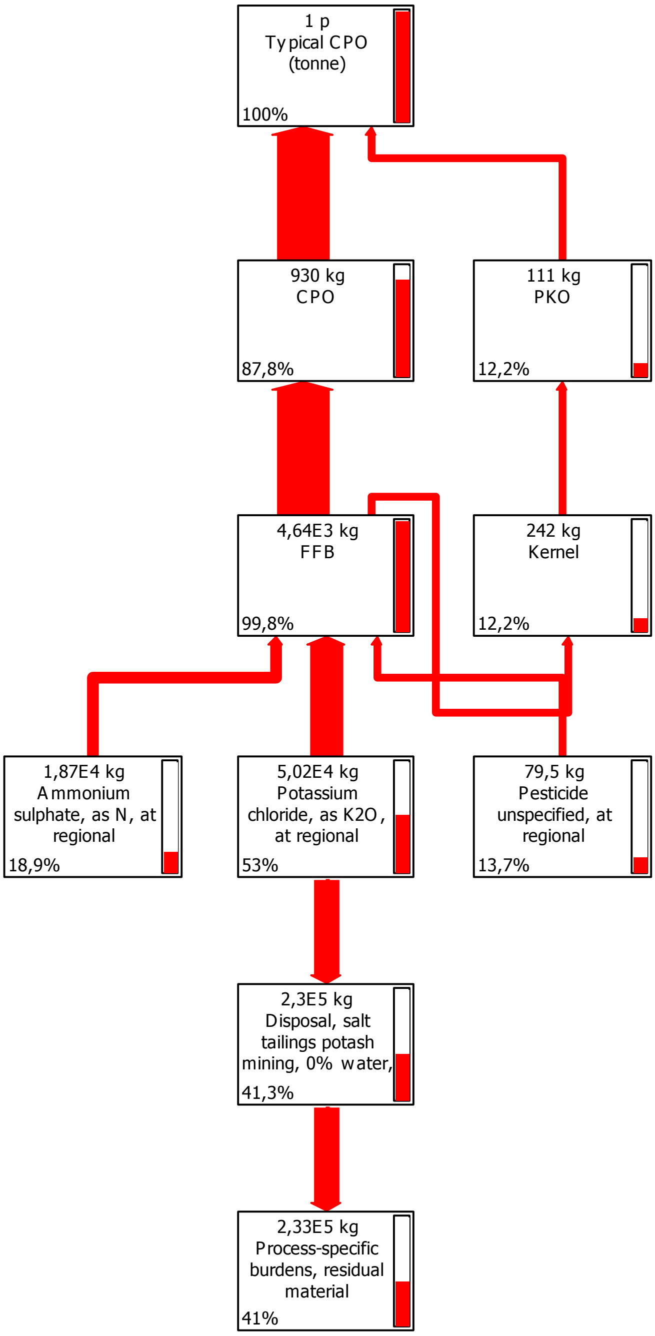


Figure A8. 1. Life-cycle tree of conventional palm oil

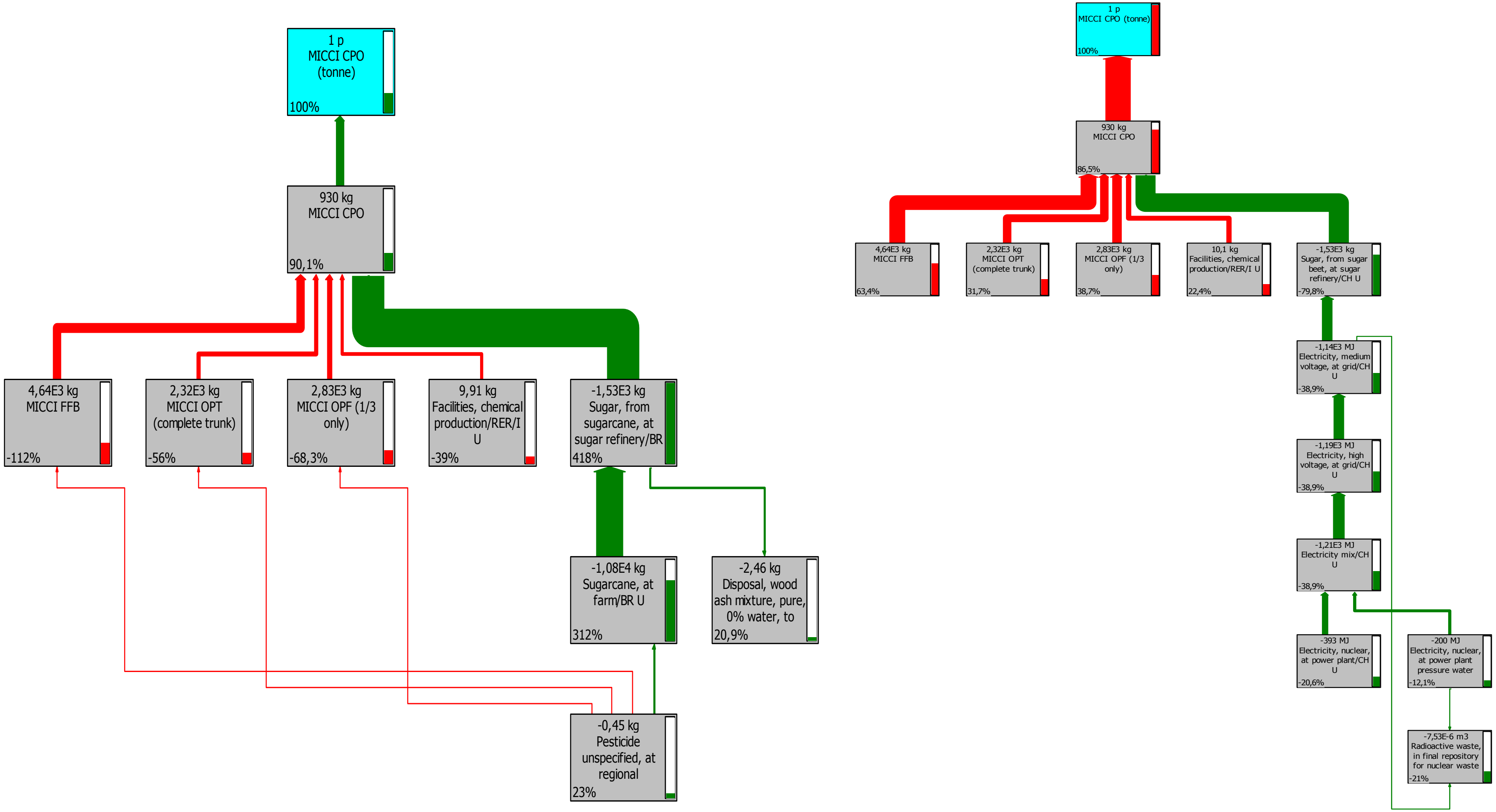


Figure A8. 2. Life-cycle tree of MICCI palm oil with FS as displacement of sugar cane (left) and sugar beet (right).
Red – Negative interventions. Green – Positive interventions

APPENDIX 9. Stream Tables

The stream tables below contain the composition and description of the process streams corresponding to the process flow diagrams in Appendix 4. The presented streams tables are based on biomass derived from three-mills and the related plantations to them.

STREAM Nr.		< 1 >		< 2 >		< 3 >		< 4 >		< 5 >		< 6 >		< 7 >	
Name		OPT		OPTA		BARK		SOPT		OPF		SOPF		PBM	
COMP	MW	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h
Acetic acid	60.05														
Biomass	-														
Ca(OH)2	74.09														
CaSO4	136.14														
Cellulose	162.14	9.88	60.95	8.60	53.04			8.60	53.04	13.76	84.86	13.76	84.86	22.36	137.91
CH4	16.04														
CO2	44.01														
Enzymes	24.02														
Extractives	-														
Fructose	180.16	0.20	1.11	0.17	0.94			0.17	0.94	0.17	0.94	0.17	0.94	0.34	1.89
Furfural	96.09														
Glucose	180.16	3.48	19.32	3.03	16.82			3.03	16.82	4.50	24.98	4.50	24.98	7.53	41.80
H2SO4	98.08														
Hemicellulose	132.12	8.25	62.44	7.18	54.34			7.18	54.34	5.41	40.95	5.41	40.95	12.59	95.29
Inerts	56.08	0.82	14.54	0.71	12.66			0.71	12.66	2.29	40.83	2.29	40.83	3.00	53.50
Lignin	-	4.10	-	3.57	-			3.57	-	5.12	-	5.12	-	8.69	-
Lignin deg.	152.15														
N2	28.01														
NO	30.00														
NH4-acetate	77.08														
O2	32.00														
Oil residues	-														
SO2	64.06														
Sulphuric ac.	98.08														
Sucrose	342.30	0.20	0.58	0.17	0.50			0.17	0.50	1.75	5.11	1.75	5.11	1.92	5.61
Water	18.02	63.00	3496.12	54.81	3041.62			54.81	3041.62	77.00	4273.03	77.00	4273.03	131.81	7314.65
Xylose	150.13	0.07	0.45	0.06	0.40			0.06	0.40					0.06	0.40
Total		90.00	-	78.30	-	11.70	-	78.30	-	110.00	-	110.00	-	188.30	-
Vol flow	m3/h	-		-		-		-		-		-		-	
Enthalpy	kW	-317673		-276376		-41298		-276376		-390147		-390147		-666523	
Phase	L/V/S	S		S		S		S		S		S		S	
Press.	Bara	1.0		1.0		1.0		1.0		1.0		1.0		1.0	
Temp	oC	25.0		25.0		25.0		25.0		25.0		25.0		25.0	

STREAM Nr	0	< 8 >		< 9 >		< 10 >		< 11 >		< 12 >		< 13 >		< 14 >	
Name	0	PBMF		SAP		EFB		SEFB		FIBRES		SBM		IMBWATER	
COMP	MW	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h
Acetic acid	60.05														
Biomass	-														
Ca(OH)2	74.09														
CaSO4	136.14														
Cellulose	162.14	22.36	137.90			4.79	29.52	4.79	29.52	5.30	32.67	32.44	200.09		
CH4	16.04														
CO2	44.01														
Enzymes	24.02														
Extractives	-					1.86	-	1.86	-	1.07	-	2.93	-		
Fructose	180.16	0.02	0.13	0.32	1.76							0.02	0.13		
Furfural	96.09														
Glucose	180.16	0.53	2.93	7.01	38.89							0.53	2.93		
H2SO4	98.08														
Hemicellulose	132.12	12.59	95.27			3.29	24.92	3.29	24.92	3.43	25.93	19.31	146.12		
Inerts	56.08	3.00	53.46			1.03	18.44	1.03	18.44	0.68	12.13	4.71	84.02	84.73	1510.89
Lignin	-	8.68	-			2.31	-	2.31	-	4.11	-	15.10	-		
Lignin deg.	152.15														
N2	28.01														
NO	30.00														
NH4-acetate	77.08														
O2	32.00														
Oil residues	-														
SO2	64.06														
Sulphuric ac.	98.08														
Sucrose	342.30	0.13	0.39	1.79	5.23							0.13	0.39		
Water	18.02	47.32	2625.82	169.22	9390.87	28.11	1559.97	28.11	1559.97	9.72	539.40	85.15	4725.18		
Xylose	150.13	0.00	0.03	0.05	0.36										
Total	0	94.63	-	178.39	178.39	41.40	-	41.40	-	24.30	-	160.33	-	84.73	1510.89
Vol flow	m3/h	-		175		-		-		-		122		85	
Enthalpy	kW	-277111		-762635		-142867		-142867		-61415		-481393		-373224	
Phase	L/V/S	S		L		S		S		S		S		L	
Press.	Bara	1.0		1.0		1.0		1.0		1.0		1.0		1.0	
Temp	oC	25.0		25.0		25.0		25.0		25.0		25.0		25.0	

STREAM	Nr.	< 15 >		< 16 >		< 17 >		< 18 >		< 19 >		< 20 >		< 21 >	
Name		HBM		SACID		DILWATER		SACID SOL		AIBM		PTTBM		PTTBM2	
COMP	MW	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h
Acetic aci	60.05											1.16	19	1.00	17
Biomass	-														
Ca(OH)2	74.09														
CaSO4	136.14														
Cellulose	162.14	32.44	200							32.44	200	30.82	190	30.82	190
CH4	16.04														
CO2	44.01														
Enzymes	24.02														
Extractive	-	2.93	-							2.93	-	2.93	-	2.93	-
Fructose	180.16	0.02	0							0.02	0	0.09	1	0.09	1
Furfural	96.09											0.40	4	0.16	2
Glucose	180.16	0.53	3							0.53	3	2.40	13	2.40	13
H2SO4	98.08													1.66	17
Hemicellu	132.12	19.31	146							19.31	146	1.27	10	1.27	10
Inerts	56.08	4.71	84							4.71	84	4.71	84	4.71	84
Lignin	-	15.10	-							15.10	-	14.95	-	14.95	-
Lignin deg	152.15											0.15	1	0.15	1
N2	28.01														
NO	30.00														
NH4-acet	77.08														
O2	32.00														
Oil residue	-														
SO2	64.06														
Sulphuric	98.08			1.66	17			1.66	17	1.66	17	1.66	17		
Sucrose	342.30	0.13	0							0.13	0				
Water	18.02	96.58	5360	0.03	2	76.45	4243	76.49	4245	173.07	9604	223.30	12392	174.268	9671
Xylose	150.13	0.00	0							0.00	0	18.56	124	18.56	124
Total	0	171.77	-	1.70	18.82	76.45	4242.69	78.15	4261.51	249.92	-	302.41	-	252.99	-
Vol flow	m3/h	141		1		77		78		220		311		224	
Enthalpy	kW	-523305		-3883		-336762		-340645		-863950		-1052678		-871378	
Phase	L/V/S	S/L		L		L		L		S/L		S/L		S/L	
Press.	Bara	1.0		1.0		1.0		1.0		1.0		12.4		1.0	
Temp	oC	100.0		25.0		25.0		25.0		71.1		190.0		100.1	

STREAM	Nr.	< 22 >		< 23 >		< 24 >		< 25 >		< 26 >		< 27 >		< 28 >	
Name		VENTA		PTTLIQ		PTTSOLID		CPTTLIQ		SLIME		WAT3		LIME	
COMP	MW	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h
Acetic aci	60.05	0.16	3	0.73	12	0.27	4	0.73	12						
Biomass	-														
Ca(OH)2	74.09									1.21	16			1.21	16
CaSO4	136.14														
Cellulose	162.14					30.82	190								
CH4	16.04														
CO2	44.01														
Enzymes	24.02														
Extractive	-			2.14	-	0.79	-	2.14	-						
Fructose	180.16			0.07	0	0.03	0	0.07	0						
Furfural	96.09	0.24	2	0.11	1	0.04	0	0.11	1						
Glucose	180.16			1.75	10	0.65	4	1.75	10						
H2SO4	98.08			1.21	12	0.45	5	1.21	12						
Hemicellu	132.12					1.27	10								
Inerts	56.08					4.71	84								
Lignin	-					14.95	-								
Lignin deg	152.15			0.11	1	0.04	0	0.11	1						
N2	28.01														
NO	30.00														
NH4-acet	77.08														
O2	32.00														
Oil residue	-														
SO2	64.06														
Sulphuric	98.08														
Sucrose	342.30					5.01	15								
Water	18.02	49.03	2721	127.29	7064	46.98	2607	127.29	7064			10.86	603	10.86	603
Xylose	150.13			13.56	90			13.56	90						
Total	0	49.43	2725.97	146.98	-	106.00	-	146.98	-	1.21	16.29	10.86	602.66	12.07	618.96
Vol flow	m3/h	84626		148		76		140		1		11		11	
Enthalpy	kW	-181300		-587026		-284352		-595874		-4463		-47854		-52317	
Phase	L/V/S	V		L				L		S		L		L	
Press.	Bara	1.0		1.0		1.0		1.0		1.0		1.0		1.0	
Temp	oC	100.1		100.1		100.1		45.2		25.0		25.0		25.0	

STREAM	Nr.	< 29 >		< 30 >		< 31 >		< 32 >		< 33 >		< 34 >		< 35 >	
Name		CONDSLUR		GYPSUM		CONDLIQ		ENZ		DILWAT2		ENZSLN		PREENZ	
COMP	MW	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h
Acetic aci	60.05	0.73	12	0.00	0	0.73	12							1.00	17
Biomass	-														
Ca(OH)2	74.09	0.29	4	0.00	0	0.29	4							0.29	4
CaSO4	136.14	1.68	12	1.68	12										
Cellulose	162.14													30.82	190
CH4	16.04														
CO2	44.01														
Enzymes	24.02							0.02	1			0.02	1	0.02	1
Extractive	-	2.14	-	0.01	-	2.14	-							2.93	-
Fructose	180.16	0.07	0			0.07	0							0.09	1
Furfural	96.09	0.11	1			0.11	1							0.16	2
Glucose	180.16	1.75	10	0.01	0	1.75	10							2.39	13
H2SO4	98.08	0.00	0											0.45	5
Hemicellu	132.12													1.27	10
Inerts	56.08													4.71	84
Lignin	-													14.95	-
Lignin deg	152.15	0.11	1			0.11	1							0.15	1
N2	28.01														
NO	30.00														
NH4-acet	77.08														
O2	32.00														
Oil residue	-														
SO2	64.06														
Sulphuric	98.08														
Sucrose	342.30													5.01	15
Water	18.02	138.59	7691	0.49	27	138.10	7664			118.69	6587	118.69	6587	303.77	16857
Xylose	150.13	13.56	90	0.05	0	13.51	90							13.51	90
Total	0	159.05	-	2.24	-	156.81	-	0.02	0.99	118.69	6586.57	118.71	6587.56	381.53	-
Vol flow	m3/h	151		1		150		0		119		119		346	
Enthalpy	kW	-648191		-7202		-640989		-21		-522811		-522832		-1448173	
Phase	L/V/S	L/S		L/S		L		S		L		L		L/S	
Press.	Bara	1.0		1.0		1.0		1.0		1.0		1.0		1.0	
Temp	oC	48.4		48.4		48.4		25.0		25.0		25.0		50.0	

STREAM	Nr.	< 36 >		< 37 >		< 38 >		< 39 >		< 40 >		< 41 >		< 42 >	
Name		ENZSLUR		LIGSOLID		LCSUG		SUG		S1		S2		S3	
COMP	MW	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h
Acetic aci	60.05	1.00	17	0.05	1	0.95	16	0.95	16	0.95	16	0.95	16	0.95	16
Biomass	-														
Ca(OH)2	74.09	0.29	4	0.02	0	0.28	4	0.28	4	0.28	4	0.28	4	0.28	4
CaSO4	136.14														
Cellulose	162.14	3.08	19	3.08	19										
CH4	16.04														
CO2	44.01														
Enzymes	24.02	0.02	1	0.02	1										
Extractive	-	2.93	-	0.15	-	2.77	-	2.77	-	2.77	-	2.77	-	2.77	-
Fructose	180.16	0.09	1	0.00	0	0.09	0	0.41	2	0.41	2	0.41	2	0.41	2
Furfural	96.09	0.16	2	0.01	0	0.15	2	0.15	2	0.15	2	0.15	2	0.15	2
Glucose	180.16	33.21	184	1.73	10	31.48	175	38.49	214	38.49	214	38.49	214	38.49	214
H2SO4	98.08	0.45	5	0.02	0	0.43	4	0.43	4	0.43	4	0.43	4	0.43	4
Hemicellu	132.12	0.13	1	0.13	1										
Inerts	56.08	4.71	84	4.71	84										
Lignin	-	14.95	-	14.95	-										
Lignin deg	152.15	0.15	1	0.01	0	0.14	1	0.14	1	0.14	1	0.14	1	0.14	1
N2	28.01														
NO	30.00														
NH4-acet	77.08														
O2	32.00														
Oil residue	-														
SO2	64.06														
Sulphuric	98.08														
Sucrose	342.30							1.79	5	1.79	5	1.79	5	1.79	5
Water	18.02	300.54	16678	15.70	871	284.84	15807	454.06	25198	454.06	25198	454.06	25198	454.06	25198
Xylose	150.13	19.81	132	1.035	7	18.78	125	18.83	125	18.83	125	18.83	125	18.83	125
Total	0	381.53	-	41.63	-	339.90	-	518.29	-	518.29	-	518.29	-	518.29	-
Vol flow	m3/h	351		34		317		492		513		516		495	
Enthalpy	kW	-1448173		-94641		-1353532		-2116167		-2094243		-2091196		-2113120	
Phase	L/V/S	L/S		L/S		L		L		L		L		L	
Press.	Bara	1.0		1.0		1.0		1.0		1.0		1.0		1.0	
Temp	oC	49.0		49.0		49.0		40.3		79.8		85.0		46.0	

STREAM Nr.		< 43 >		< 44 >		< 45 >		< 46 >		< 47 >		< 48 >		< 49 >	
Name		PSUG		VENTB		POME		POMEB		WWAT		TTWATC		BIOGAS	
COMP	MW	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h
Acetic aci	60.05	0.95	16	0.16	3					0.16	3	0.02	0		
Biomass	-											0.01	-		
Ca(OH)2	74.09	0.28	4												
CaSO4	136.14														
Cellulose	162.14														
CH4	16.04													0.59	37
CO2	44.01											0.04	1	1.31	30
Enzymes	24.02														
Extractive	-	2.77	-			1.23	-	1.23	-	1.23	-	0.12	-		
Fructose	180.16	0.41	2												
Furfural	96.09	0.15	2	0.24	2					0.24	2	0.02	0		
Glucose	180.16	38.49	214												
H2SO4	98.08	0.43	4												
Hemicellu	132.12														
Inerts	56.08														
Lignin	-														
Lignin deg	152.15	0.14	1												
N2	28.01														
NO	30.00														
NH4-acet	77.08														
O2	32.00														
Oil residu	-					0.25	-	0.25	-	0.25	-	0.03	-		
SO2	64.06														
Sulphuric	98.08														
Sucrose	342.30	1.79	5												
Water	18.02	454.06	25198	49.03	2721					49.03	2721	87.57	4860		
Xylose	150.13	18.83	125			38.88	259	38.88	259	38.88	259			0.11	1
Total	0	518.29	-	49.43	2725.97	40.38	-	40.38	-	89.81	-	87.80	-	2.00	67.10
Vol flow	m3/h	485		54		42		41		94		90		1899	
Enthalpy	kW	-2124244		-215634		-171469		-173280		-388912		-384514		-4398	
Phase	L/V/S	L		L		L		L		L		L		V	
Press.	Bara	1.0		1.0		1.0		1.0		1.0		1.0		1.0	
Temp	oC	25.0		40.0		80.0		40.0		40.0		42.0		42.0	

STREAM	Nr.	< 50 >		< 51 >		< 52 >		< 53 >		< 54 >		< 55 >		< 56 >	
Name		LAGAIR		TTWATE		LOSSES		TTPOMEF		SLUDGE		BFW		BFWB	
COMP	MW	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h
Acetic aci	60.05														
Biomass	-			0.14	-					0.14	-				
Ca(OH)2	74.09														
CaSO4	136.14														
Cellulose	162.14														
CH4	16.04														
CO2	44.01			0.00	0	0.27	6	0.00	0						
Enzymes	24.02														
Extractive	-			0.00	-			0.00	-						
Fructose	180.16														
Furfural	96.09														
Glucose	180.16														
H2SO4	98.08														
Hemicellul	132.12														
Inerts	56.08														
Lignin	-														
Lignin deg	152.15														
N2	28.01	7.68	274	0.00	0	7.68	274	0.00	0						
NO	30.00														
NH4-acet	77.08														
O2	32.00	2.04	64			1.85	58								
Oil residu	-			0.00	-			0.00	-						
SO2	64.06														
Sulphuric	98.08														
Sucrose	342.30														
Water	18.02	0.28	16	87.65	4864	0.20	11	86.77	4815	0.88	49	129.72	7199	129.72	7199
Xylose	150.13														
Total	0	10.00	353.48	87.80	-	10.00	349.19	86.79	-	1.02	-	129.72	7198.67	129.72	7198.67
Vol flow	m3/h	8650		88		8656		87		1		131		131	
Enthalpy	kW	-1099		-386562		-1408		-382244		-4318		-571392		-569876	
Phase	L/V/S	V		L		V		L		L/S		L		L	
Press.	Bara	1.0		1.0		1.0		1.0		1.0		1.0		1.0	
Temp	oC	25.0		25.0		25.0		25.0		25.0		25.0		37.6	

STREAM	Nr.	< 57 >		< 58 >		< 59 >		< 60 >		< 61 >		< 62 >		< 63 >	
Name		BFWC		COMBAIR		COMBSOL		CFLUEG		SHSTM		MPSTMA-PTT		LPSTM	
COMP	MW	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h
Acetic aci	60.05														
Biomass	-														
Ca(OH)2	74.09														
CaSO4	136.14														
Cellulose	162.14														
CH4	16.04														
CO2	44.01							47.30	1075						
Enzymes	24.02														
Extractive	-														
Fructose	180.16														
Furfural	96.09														
Glucose	180.16														
H2SO4	98.08														
Hemicellu	132.12														
Inerts	56.08					4.71	84								
Lignin	-														
Lignin deg	152.15														
N2	28.01							156.42	5584						
NO	30.00							0.02	1						
NH4-acet	77.08														
O2	32.00			41.54	1298			5.34	167						
Oil residue	-														
SO2	64.06							0.02	0						
Sulphuric	98.08			156.42	1595										
Sucrose	342.30														
Water	18.02	129.72	7199	5.70	316			34.51	1915	134.38	7457	52.50	2913	81.88	4544
Xylose	150.13														
Total	0	129.72	7198.67	203.66	3209.32	4.71	84.00	243.60	8741.57	134.38	7457.27	52.50	2913.21	81.88	4544.06
Vol flow	m3/h	131		176186		1		830824		7150		10637		563008	
Enthalpy	kW	-561031		-22374		-13801		-251487		-469520		-188728		-300206	
Phase	L/V/S	L		V		S		V		V		V		V	
Press.	Bara	1.0		1.0		1.0		1.0		63.0		12.0		2.8	
Temp	oC	95.0		25.0		870.0		55.0		454.0		276.0		144.0	

STREAM	Nr.	< 64 >		< 65 >		< 66 >		< 67 >		< 68 >	
Name		LPSTMC-PAST		LPSTMP-OUT		STMCON		BFWD		LPSTMA-PTT	
COMP	MW	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h	tonne/h	kmol/h
Acetic aci	60.05										
Biomass	-										
Ca(OH)2	74.09										
CaSO4	136.14										
Cellulose	162.14										
CH4	16.04										
CO2	44.01										
Enzymes	24.02										
Extractive	-										
Fructose	180.16										
Furfural	96.09										
Glucose	180.16										
H2SO4	98.08										
Hemicellu	132.12										
Inerts	56.08										
Lignin	-										
Lignin deg	152.15										
N2	28.01										
NO	30.00										
NH4-acet	77.08										
O2	32.00										
Oil residue	-										
SO2	64.06										
Sulphuric	98.08										
Sucrose	342.30										
Water	18.02	4.66	259	65.79	3651	4.66	259	134.38	7457	11.43	635
Xylose	150.13										
Total	0	4.66	258.60	65.79	3650.94	4.66	258.60	134.38	7457.26	11.43	634.52
Vol flow	m3/h	3204		45243		5		146		7861	
Enthalpy	kW	-17087		-241210		-20133		-581164		-41913	
Phase	L/V/S	V		V		L		L		V	
Press.	Bara	2.8		2.8		1.0		1.0		2.8	
Temp	oC	144.0		144.0		99.0		95.1		144.0	

APPENDIX 10. References

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