PROCESS MODELLING OF SUGAR MILL BIOMASS TO ENERGY CONVERSION PROCESSES AND ENERGY INTEGRATION OF PYROLYSIS

By

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Abstract

The sugar industry over the years has been producing sugarcane bagasse as part of the sugar milling process. Currently this sugar mill biomass is incinerated inefficiently as a means of their disposal to produce steam and electricity, which in most cases are only just enough to supply the energy required to run the mills, thereby leaving very little or no extra energy for sale to bring in extra income in addition to sales revenue from sugar. However, the recent instability and uncertainties in the price of sugar and the global call for a green and sustainable environment have necessitated the search for ways of making effective use of this biomass to supply sugar mill energy demands, while producing extra energy in the form of electricity and other energy products for sale and at the same time contributing towards environmental sustainability.

The main objective of this work was to develop process models for the processing of sugar mill biomass into energy and energy products. Based on this, biomass to energy conversion process (BMECP) models have been developed for various process configurations of two thermochemical processes; Combustion and Fast Pyrolysis using the Aspen Plus® simulation software. The aim of process modelling was to utilizing sugar cane bagasse as an input energy source to supply the energy requirements of two sugar mill configurations (efficient and less efficient mills), while generating extra electricity and high valued energy products for sale. Four BMECP configurations; 30bar BPST, 40bar CEST, 63bar CEST and 82bar CEST systems were modelled for the combustion thermochemical process. For the fast pyrolysis thermochemical process, two process configurations: Pure Fast Pyrolysis BMECP and Partial Fast Pyrolysis BMECP were modelled. The former BMECP utilizes all available bagasse through fast pyrolysis to produce bio-oil and biochar alongside generating electricity as well as energy to run the sugar mill operations. In the latter BMECP model, only surplus bagasse after separation of the quantity needed to supply the sugar mill energy requirement and electricity production is used to produce bio-oil and biochar.

The technical performance of the BMECP models have been analysed and compared based on steam and electricity production rates, process efficiencies and environmental impacts (based on CO_2 savings). The effects of boiler operating pressure and bagasse moisture content on the performance of the combustion based BMECP models have also been investigated. Finally, detailed economic models have been developed using the Aspen Process Economic Analyzer (Icarus®) to assess the economic viability of the BMECP models and sensitivity analysis performed to study the response of the BMECP models to variations in economic parameters. Technical performance analysis shows the combustion based BMECP models perform better than the Pure Fast Pyrolysis and Partial Fast Pyrolysis BMECP models with regards to steam and electricity production, thereby giving them higher electrical efficiencies. The electricity generation rate has been shown to increase with increasing boiler operating pressure and decreasing bagasse moisture content while steam production rate has been shown to increase with decreasing bagasse moisture content and decreasing boiler operating pressure. Despite the lower electrical efficiencies of the fast pyrolysis based BMECP models, the analysis shows that their overall process efficiencies compare very well with those of the combustion based BMECP models due to the production of high energy value pyrolysis products. Based on common operating pressure and 50% bagasse moisture content, the Pure Fast Pyrolysis and the Partial Fast Pyrolysis models have proved to be more environmental friendly with hourly CO_2 savings of 40.44 and 41.30 tons for the Partial Fast Pyrolysis BMECP and the Pure Fast Pyrolysis BMECP respectively based on a 300 ton of sugarcane/h (81 ton bagasse/h) plant size.

From an economic point of view, biomass combustion based on the 63bar CEST BMECP model has proved to be the most economically viable option under current economic conditions. First order total capital investment estimate for this BMECP is about $116 million, producing NPV of $390 million at the end of a 20 year plant life and IRR of 34.51%. The Pure Fast Pyrolysis BMECP model is the least economic viable option. Sensitivity analysis shows this BMECP model is the most sensitive to changes in bagasse and electricity prices; recording -191.61/+446.86% change in NPV for a ±30% change in bagasse price and -91.5/+338.60% for a ±30% change in electricity price.

Opsomming

Die afgelope jare het suikerriet-afval (bagasse) by suikermeule ‘n belangrik byproduk van die suiker-industie geraak. Tans word hierdie afval of biomasse verbrand in die suikermeule se poging om stoom en elektrisiteit op te wek; maar die die proses is oneffektief. Die hoeveelheid energie wat opgewek word, is skaars genoeg om die suikermeule self aan die gang te hou; daar is feilik geen sprake ‘n surplus energie waaruit ekstra inkomste verkry kan word toevoegend tot inkomste uit die suiker verkope self. Die huidige onstabiele suikerprys en gepaardgaande onsekerhede sowel as die werêldwye oproep vir ‘n groen- en volhoubare omgewing, noodsaak ‘n nuwe soeke na effektiewe manier om die afvalmateriaal sinvol te verwerk. Die tipe effektiwiteit van verwerking waarna gesoek word moet die volgende uitkomste hê: verskaffing van genoeg energie tydens produksie aan die suikermeuele self; vervaardiging van ekstra energie in die vorm van eletrisieteit en ander energie produkte. Terselfder moet die ook bydra tot die volhoubaarheid van die omgewing. Die grootste gedeelte van hierdie navorsing is gewy aan die ontwikkeling van “proses modelle” om suikemeule afval (bagasse) te omskep in energie en energie-produkte. Om hierdie doel te bereik, is biomassa-tot-energie omskeppingsproses- modelle (BMECP) ontwikkel om verskeie proses konfigurasies van twee termo-chemiese prosesse, naamlik Verbranding (Combustion), en Vinnige Pirolise (Fast Pyrolysis) deur die gebruik van die ‘Aspen Plus®’- simulasie sagteware.

Die doel van die proses modelering was om suikerriet biomassa as ‘n bron van energie te gebruik om weer die energie benodighehede van twee denkbeeldige suikermeule vas te stel; een meul is voorgestel as effektief, die ander as minder effektief. Terselfdertyd is gekyk na die hoeveelheid ekstra energie wat elkeen sou opwek en ander hoogs waardevolle energie produkte om te verkoop (bv. ‘bio-olies en bio-char’). Vier “BMECP” konfigurasies (voorstellings) 30bar BPST, 40bar CEST, 63bar CEST en 82bar CEST sisteme is gemodelleer vir die Verbranding termo-chemiese proses. In die geval van die Pirolise (Pyrolysis) termo-chemiese proses, is twee proses konfigurasies gemodelleer: 1. Suiwer Vinnige Pyrolyise BMECP en 2. Gedeeltelik Vinnige Pirolise BMECP. In die geval van eersgenoemde, word alle beskikbare ‘bagasse’ deur vinnige pirolise omskep om ‘bio-olie’ en ‘bio-char’ te vervaardig.Verder wek dit ook elektrisiteit op so wel as die nodige energie om die suikermeule te laat opereer. In die geval van die Gedeeltlike Vinnige Pirolise BMECP , moet daar eers genoegsame ‘bagasse’ opsy gesit word om die suikermeule van genoegsame energie te voorsien vir die volle funskionering daarvan en elektrisiteit-opwekking. Van die surplus of oorblywende ‘bagasse’ kan dan gebruik word om ‘bio-olie’ en ‘biochar’ te produseer.

Die tegniese prestasie van al die BMECP modelle is geanaliseer en vergelyk ten opsigte van stoom en elektrisiteits-opwekking; proses effektiewiteit asook die impak op die omgewing (gebaseer op CO_2 – besparings). Die effek van stoomkettel-druk tydens operering asook die bagasse se vog-inhoud. Op die prestasie van die verbrandingsgebaseerde modelle is ook ondersoek. Laastens, uitgebreide ekonomeidese modelle is ook ontwikkel deur die gebruik van die ‘Aspen Process Economic Analyser (Icarus®)’. Sodoende is die ekonomiese vatbaarheid van die BMECP modelle ondersoek. Hierdie sagteware help ook met. Sensitiwiteits-analise in die bestudering van die terugvoer van die BMECP modelle tot veranderlikes in ekonomiese parameters.

Rakende effektiwiteit, toon die uitslae dat die verbrandings-gebaseerde BMECP modelle beter vaar as die met betrekking tot stoom- en elektrisiteits-opwekking. Verbrandings-gebaseerde-modelle toon hoër elektriese effektiwiteit. Indien die vog-inhoud van die bagasse laag was en die tempo van stoomketel operasie druk verhoog is, het die tempo van elektriesiteits-opwekking ook gestyg. Ten opsigte van stoom daarenteen, het die stoom-opwekking tempo verhoog in die die vogl inhou van diebagasse laag was asook verminderde stoomketel operering druk. Ten spyte van die laer elektriese effektiewiteit van die Suiwer Vinnig- en Gedeeltelik Vinnig BMECP modelle, dui die analise aan dat hul proses effektiewiteit in die geheel Goed vergelyk met die van die verbrandings-gebaseerde BMECP modelle. Dit is toe te skryf aan die produksie van die hoë-energie draende pirolise produkte. Gebaseer op algemene operering druk van 50% ‘bagasse’ vog-inhoud, het die bogenoemde twee modelle bewys om meer omgewings-vriendelik te wees met uurlikse CO_2-besparings. In die geval van Gedeeltelike Vinnige Pirolise BMECP, 40.44 en vir die Suiwer Vinnige Pirolise BMECP 41.30 gebaseer op ‘n 300 ton suikerriet/h (81 ton bagasse/h) plantasie-grote.

Ten slotte, vanuit ‘n ekonomiese oogpunt, blyk ‘n biomassa verbranding gebaseer op die 63 bar CEST BMECP model die mees ekonomies-vatbare opsie onder huidige ekonomiese omstandighede. Eerste orde totale kapitale belegging beraming vir hierdie BMECP is ongeveer $116 miljoen, produksie NPV is $390 miljoen aan die einde van ‘n 20 jaar tydperk vir ‘n suikerriet-aanleg. IRP is 34.51%. Die Suiwer Vinnige Pirolise BMECP is die mins-ekonomiese vatbare model. Sensitiewiteits-analises het getoon dat hierdie BMECP model baie sensitief is ten opsigte van verandering in die pryse van bagasse en elektrisieteit; in die geval van NPV is veranderinge van -191.61/+446.86% aangedui op ‘n ±30% verandering in bagasse pryse. In die geval van elektrisieteitspryse, is ‘n sensitiewiteit van van -91.5/+338.60% op ‘n ±30% prysverandering getoon.

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Dedication

To my parents for their magnificent love and support.

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Nomenclature

$/MW	US dollar per megawatt
bar	10^5 Pascal
BPST	Back Pressure Steam Turbine
C_{at}	adjusted total capital cost
CEST	Condensing Extraction Steam Turbine
CHP	Combined Heat and Power
C_t	total capital cost
e	project capital escalation
$E_{elec.}$	Net electric power output
E_{th}	thermal energy in feed/products
gal	gallon
GJ	gigajoules
H	weight % hydrogen
HHV	Higher Heating Value
IRR	Internal Rate of Return
kW/tch	kilo Watt per ton of cane crushed
kWh	kilo Watts-hour
LCOE	Levelised Cost of Electricity
LHV	Lower Heating Value
LHV_{ar}	LHV of bagasse as received
LHV_{dry}	LHV of dry bagasse
MJ	megajoule
MW_e	megawatt electricity
MW_{th}	megawatt thermal energy
n	exponential factor
NPV	Net Present Value
PI	Profitability Index
PO	Payout period
tch	tons of cane crushed per hour
tons	tonnes (10^3 kg)
x	moisture content as received
α	conversion efficiency
$\eta_{electrical}$	electrical efficiency
$\eta_{overall}$	overall process efficiency

CHAPTER ONE

1.1 Introduction

The increase in the demand for energy caused by the increase in global industrialization, the rapid rate at which fossil oil reserves are depleting, as well as issues of environmental concern with regards to greenhouse gas emissions, have encouraged the search for alternative energy sources, mainly from renewable resources such as biomass (Goyal et al., 2008; Dias et al., 2009; Nguyen et al., 2009; Lu et al., 2009; Garcia-Perez, 2010; Venderbosch and Prins, 2010). Biomass provides a clean and renewable source of energy. Converting biomass to energy rich products is CO_2 neutral as any CO_2 produced during the conversion process is reabsorbed from the atmosphere by plants (Basu, 2010). Also the emission level of NO_x and SO_x from biomass compared to that of fossil based fuels is almost zero since biomass contains very low percentages of N and S (Nikoo & Mahinpey, 2008). Biomass has been successfully converted to energy sources such as heat, electricity and even fuel-grade oils through both thermochemical and biological processes (Bridgwater, 2011; Bridgwater, 2003; Bridgwater, 2001). The use of biomass as an energy source however depends very much on biomass availability.

Sugarcane bagasse is one such source of readily available biomass. Bagasse is the fibrous material that remains after juice is extracted from sugarcane during the sugar manufacturing process, and like any other biomass, it is made up mainly of cellulose, hemicelluloses, lignin and some small fraction of extractives (Howard et al., 2003; Tsai et al., 2006; Dias et al., 2009; Saxena et al, 2009; Venderbosch and Prins, 2010; Sluiter et al., 2010). Sugar production from sugarcane remains as one of the predominant agro-industrial activities in South Africa, producing sugar as the main product and in some instances excess of electricity after meeting the industry’s energy demand. A substantial amount of bagasse is generated in this industry during the milling process (270kg bagasse/ton of cane milled) according to Garcia-Perez et al. (2002). In South Africa, about 297kg bagasse/ton of cane was generated by the sugar industry during the 2010/2011 milling season (S. Davis, SMRI, personal communication). Currently, this waste is inefficiently combusted as solid fuel in cogeneration systems attached to most sugar mills around the world to raise steam which is then used to provide the thermal and electrical energy demand of the industry (Mbohwa, 2003; Pippo et al.,2007; Ensinas et al., 2009). Very little or no surplus bagasse is made available due to the energy intensive nature of the sugar manufacturing process and the relatively low efficiencies of both the cogeneration systems and the production process with regards to the use of energy (Ensinas et al., 2007a).

Given the rapidly changing market for sugar and the instability and uncertainties in the price of sugar, it has become important for sugar factories to introduce some form of product diversification in the industry (Ogden et al., 1990; Banerjee et al., 2003). The production of valuable products from bagasse is one way in which sugar factories can bring in added benefits. Bagasse has significant potential as energy source, which has not been fully exploited by the sugar industry (Pippo et al., 2007). Among the diversification that can be introduced into the sugar industry are the generation of excess power through improvement in efficiency of biomass combustion process and the production of fuel and speciality chemicals from bagasse. Exploring the potential of bagasse, however, requires the availability of a sufficient amount of bagasse and this in turn calls for improvement in process efficiencies and the optimal use of energy in sugar mills.

Energy integration in the sugar industry has been identified as a way of minimising the waste of energy and ensuring the proper use of energy (Ensinas et al., 2007b; Ensinas et al., 2009). The implementation of energy integration measures within the sugarcane milling process itself will thus make sufficient bagasse available, since the external thermal energy demand of the mill will be reduced drastically, implying less bagasse needed for steam generation. However, storing large quantities of bagasse for future use is not beneficial to the sugar industry in financial terms. Bagasse has a low bulk density (Bridgwater et al., 1999; Pippo et al., 2007; Pippo et al., 2009), hence requiring large volume for storage, which is very expensive. Moreover, stockpiling bagasse and other sugarcane residues poses an environmental threat to sugar mills and their surroundings because bagasse is self-combustible and may spontaneously combust if stockpiled for longer periods (Lavarack, et al., 2002; Pippo et al., 2007). This means that bagasse must be readily converted to valuable energy sources such as electricity in highly efficient cogeneration systems for sale to the grid as is done in Mauritius and Reunion (Mbohwa, 2003). The one-time use of bagasse implies that the sugar mills will have to depend on fossil based fuel for energy generation during off-season, and to avoid this, the need arises to search for alternative ways of converting bagasse and other sugarcane agro-industry waste into products that can easily be stored for future use, including pyrolysis products such as charcoal and bio-oil.

One way of converting bagasse into storable product is by the use of pyrolysis (Pippo et al., 2009; Bridgwater, 2003; Bridgwater, 2011). Generating energy products from waste biomass obtained from the sugarcane harvesting/milling process through pyrolysis has become economical and environmentally interesting. Pyrolysis, a thermochemical process, has been used to convert biomass such as bagasse into products (bio-oil and char) with a high energy density (Tsai et al., 2006). Unlike other thermochemical processes such as gasification and combustion where the syngas and heat generated, respectively, have to be used readily on site, the products of pyrolysis can be stored and used later when the need arises (Pippo et al., 2009; Bridgwater, 2003; Bridgwater, 2011). The bio-oil and biochar produced can be used for electricity production during both in-season and off-season (Pippo et al., 2009), hence ensuring all year round electricity production of which surplus can be offered for sale to the grid to generate extra income for the sugar industry. Also char can be upgraded to activated carbon which can be used in the sugar refinery process to remove colour (Devnarain et al., 2002). Char can also be used as soil amendment agent/soil additive alongside fertilizers on sugarcane plantations to improve the fertility of the soil (Brown et al., 2011; Carrier et al., 2012) which subsequently will lead to increased sugar cane yields. Studies have shown that soils that receive a combined application of fertilizer and char exhibit better plant growth resulting in yields of as high as 50% over and above that which can be obtained from soils that are given only fertilizer (Steiner et al., 2007; Tenenbaum, 2009). Apart from these benefits, pyrolysis also has the ability to supply the thermal and electrical energy needed for the sugarcane milling/sugar production process especially in the case of fast pyrolysis. Due to the high temperatures at which the technology of fast pyrolysis operates, as much energy as possible can be harnessed in the form of high pressure steam during pyrolysis products recovery, which can then be used to provide the thermal and electric energy duty of the sugar mill plant.

Hence considering the high energy demand of the sugar mill and the high quantity of thermal energy that can be recovered from the pyrolysis plant, the introduction of pyrolysis in a sugar mill through the implementation of efficient and effective energy integration networks seems to be a better technology that need to be embraced. In this way the sugar industry can benefit from producing valuable products (bio-oil and char) from fast pyrolysis, while also meeting its thermal and electrical needs from the heat recovered from the pyrolysis plant and even generating surplus electricity for sale.

This work therefore seeks to develop process models (using Aspen Plus® simulation software) for the efficient conversion of sugar mill biomass to energy (steam and electricity) and/or energy products. Notably, models will be developed for combustion (the current technology used in the sugar industry) and pyrolysis process technologies, with the aim of investigating the possible introduction of the bio-refinery concept into the sugar mill to convert sugar mill biomass (bagasse) into energy dense products while also meeting the electricity and steam demand of the mill. Models developed will be assessed to determine their capacity to provide the required process steam and electricity for the sugar mill which will be based on various process energy integration scenarios adapted in the milling process. The environmental impact (in terms of CO_2 savings) of all process model scenarios will be analysed in order to determine their contribution towards a reduction in global warming. Efficiencies of all process models will also be estimated, which will then allow for a comparison of various process technologies. The economic viability of processes will also be evaluated in order to assess the impact of these on the overall economics of the sugar industry.

1.2 Research Proposal

1.2.1 Motivation

The sugar industry, an age-old industry, is a highly energy intensive industry requiring a sufficient amount of energy (thermal and electrical) for its manufacturing process. Improvement in process efficiency has been identified as a way of cutting down on energy demand and making sufficient bagasse available, a by-product which can readily be converted into other valuable products. Currently, bagasse is combusted inefficiently in mills to cogenerate steam and electricity, and improvement in the cogeneration system is required to ensure the production of surplus energy that will generate an extra source of income for the sugar industry. Pyrolysis (fast, vacuum and slow), a thermochemical process, aside from converting biomass into highly energy dense products have the capability of generating sufficient energy in the form of heat which can be recovered to raise steam and produce electricity to meet the energy requirements of sugar mills. Though pyrolysis had been in existence for the past three decades and has successfully been applied to convert sugarcane bagasse and other biomass feeds into valuable products such as liquid fuel, its application has always been as a stand-alone process and the implementation of the technology as an integrated part of the sugar mill is under developed. Little information exists in literature with regards to pyrolysis implementation in the sugar mill and there is the need therefore for much work to be done in this field of research to help the sugar industry.

1.2.2 Objectives

The main objective of this work was to develop process models of process technologies, specifically combustion and pyrolysis, to convert sugar mill biomass into useful energy/energy products, and to compare the two in terms of efficiency and capability of meeting the energy demands of the sugar mill. The latter comparison include an assessment of the economic viability of the process technologies and their impact on the overall economics of the sugar mill especially in the context of South Africa, based on models developed.

Though the Sugars^{TM} software program is used currently to model the operations of sugar mills, its application when it comes to the concept of bio-refinery is limited, hence the need for more powerful simulation softwares such as Aspen Plus$^{®}$. Aspen Plus$^{®}$ permits the use of energy integration network tools to provide more optimal energy use scenarios for existing industrial processes, together with a business case for capital investment required for such process improvement.

Thus to achieve the above mentioned objective, different process model scenarios were developed using the Aspen Plus$^{®}$ simulation software (Aspen Technology, Inc.) to generate mass and energy balances for process flow streams and equipment of various flow sheet configurations. The mass and energy balances were then used to estimate and compare the efficiencies of different process models, including greenhouse gas emissions. Process models together with their respective mass and energy flows were then imported into Aspen Icarus® (Aspen Technology, Inc.) for the economic analysis.

The specific objectives of this work were therefore as follows:

i. To develop an Aspen Plus® model of the existing biomass to energy conversion process (combustion to steam and electricity production) of a typical sugar mill focusing on the maximum recovery of energy to meet the demands of the industry.

ii. To develop an Aspen Plus® model of the pyrolysis process based on experimental data for the pyrolysis of bagasse and analyse its capability to generate enough steam and electricity to meet the demands of the mill.

iii. Assessment and comparison of the overall process energy efficiencies of the two technologies above as well as levels of greenhouse gas emissions.

iv. To carry out an economic assessment of each of the scenarios each scenario analysed using Aspen Icarus®. This will help in decision making as to the implementation of the pyrolysis bio-refinery concept in the sugar mill.

1.2.3 Thesis Layout

Figure 1 is a block flow diagram of the thesis layout and it illustrates how the different chapters are integrated into the thesis. The first chapter gives a general introduction to the thesis and also spells out the objectives of the study. Chapter two is the literature study and it presents a general overview of the South African sugar industry as well as a general description of the sugar making process. This chapter also discusses the sugar mill biomass (bagasse) giving information on its composition and properties and their relevance to combustion and pyrolysis applications. Literature studies on combustion and pyrolysis as technologies for the conversion of biomass into energy and energy products are also contained in chapter two.

The design and approach section (Chapter three) gives a general overview of the Biomass to Energy Conversion Processes (BMECP) modelled in this study. It provides information regarding the choice of scenarios as well as the assumptions made in this work in building the process models in Aspen Plus® to generate the mass and energy balances of the various process model scenarios. Chapter three also explain the formulae and methods used in this work to estimate the process energy efficiencies as well as the savings in CO_2/environmental impacts of process models developed.

In chapter four, detailed description of the various BMECP models as implemented in Aspen Plus® is given. The results from the process simulation and the analysis, discussion and interpretation thereof are also given in this chapter. The discussion is more focused on the quantity of steam and electricity generated by each BMECP technology as well as the conversion efficiencies and environmental impacts associated with them. A comparison of results from different processes is also presented in chapter four.

Chapter five discusses the assumptions and the procedure followed in Aspen (Icarus®) to develop the economic models for the BMECP plants modelled in Aspen Plus®. Results obtained are analysed to establish the economic viability of the different BMECP technologies. The sensitivity of these models to changes in market conditions such as changes in raw material and product prices is also given in this chapter.

Chapter six is the concluding chapter and this is where the main findings from the study are presented. Recommendations for further studies are also given in this chapter.

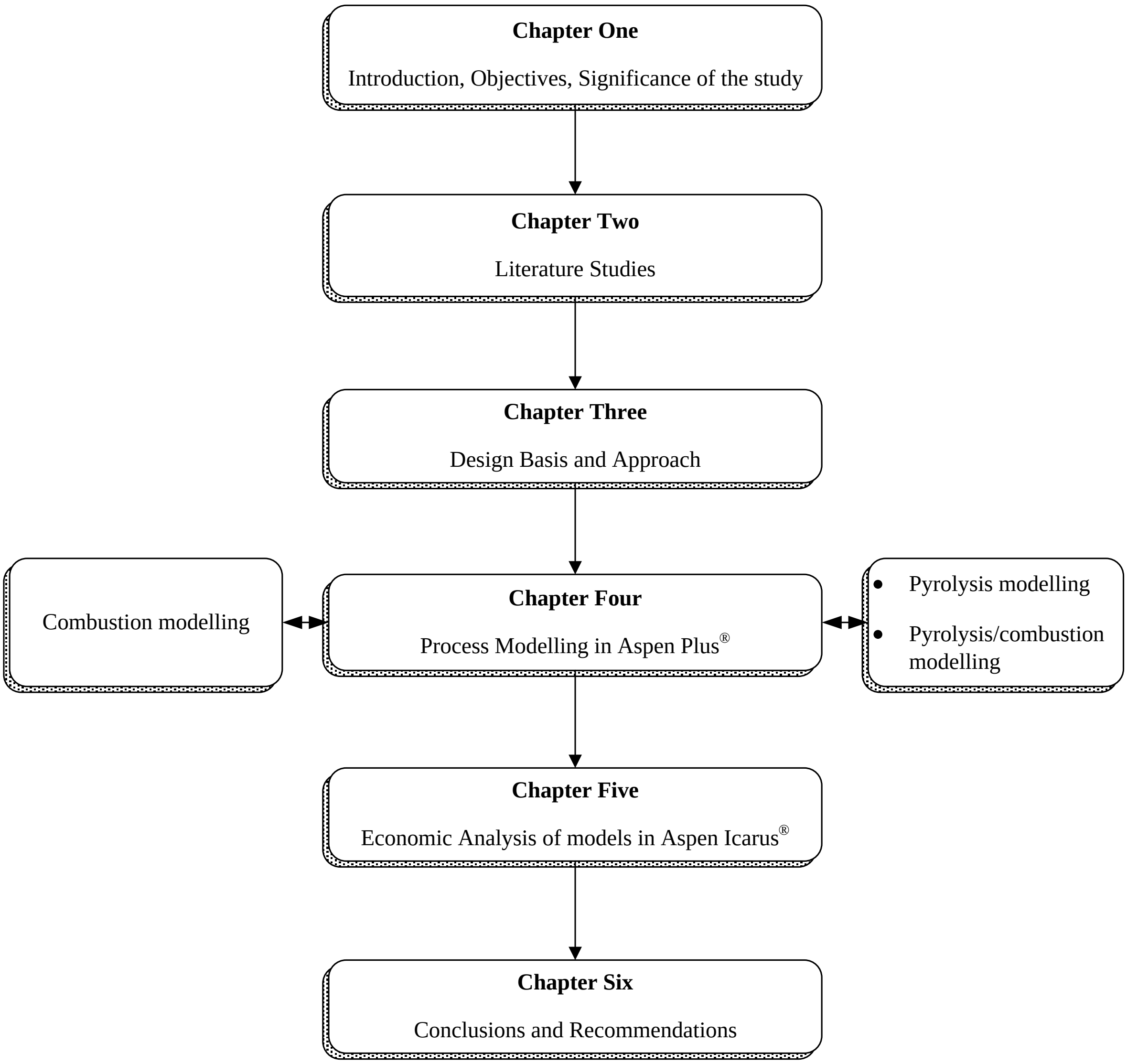

Figure 1: Flow diagram of thesis layout

1.2.4 Impacts of the Study

This research will be of particular importance to the South African sugar industry as it will bring diversification in the industry with regards to the use of sugarcane. Additional revenue will be generated for the industry through the sale of excess electricity and/or pyrolysis products. In a broader sense the bio-oil produced could be used to replace fossil fuels and this will cut down on net emissions to the environment resulting in reductions in global warming. Moreover, the products of pyrolysis can be stored and used in the production of electricity during the off-season operation periods of the sugar mill, hence ensuring an all year round generation of electricity for sale to the grid.

CHAPTER TWO

Literature

2.1 The South African Sugar Industry

2.1.1 Overview

The sugar industry in South Africa is among the top producers of sugar in the world. It specialises in the production of both raw and refined sugar alongside other by-products. The current production capacity of the industry for a season stands at about 2.5 million tons of sugar, of which about half is used locally and the rest is sold on the international market. A significant contribution to the economy of South Africa is made by the industry through the export of sugar. An average of R2.38 billion is estimated to be the industry’s contribution to the country’s foreign exchange (South African Sugar Industry Directory, 2011/2012).

Cane production is limited to the KwaZulu-Natal, Mpumalanga and Eastern Cape provinces and there are about 50000 sugarcane growers in these provincial areas registered with the South African Cane Producers Association. Most of these are small scale producers. Besides these cane growers, large plantations are also owned by most sugar mills. On average about 22 million tons of sugarcane is delivered to the mills annually (www.sasa.org.za).

There are currently 14 sugar mills in South Africa; five of these own and operate their own sugar refineries. It is estimated that these mills employ about 11,000 people (South African Sugar Industry Directory, 2011/2012).

2.1.2 Production Statistics

The average annual sugarcane production in South Africa is currently about 22 million tons from which about 2.5 million tons of sugar is recovered by the mills every season. The production statistics for the seasons 1994/5 to 2009/10 are shown in Table 1. Sugarcane yields have reduced during the last few years, due to drought conditions, thereby placing the industry under additional financial pressure, providing further motivation for the development of biorefineries. The high increase in electricity prices has contributed further to the financial pressure especially on growers using irrigation.

Table 1: South Africa Sugarcane Production Statistics

Season	Production (x 10^6), tons
1994/95	15.683
1995/96	16.713
1996/97	20.950
1997/8	22.154
1998/9	22.930
1999/00	21.223
2000/01	23.876
2001/02	21.156
2002/03	23.012
2003/04	20.418
2004/05	19.094
2005/06	21.052
2006/07	20.278
2007/08	19.723
2008/09	19.255
2009/10	18.655

Sources: Smith et al. (2010); SASA (taken from Sugar Outlook, April, 2009).

2.2 The Cane Sugar Production Process

The production of sugar from sugarcane is done in several processing steps: cane preparation and juice extraction, juice treatment and clarification, juice evaporation, sugar boiling/ crystallization, centrifugation and drying (See Figure 2). These processing steps are described below. The descriptions follow directly the works of Ensinas et al. (2009), Ensinas et al. (2007a) and Ensinas et al. (2007b) unless otherwise stated.

2.2.1 Cane Preparation and Juice Extraction

Cane preparation and juice extraction are the first steps in the sugar manufacturing process. Ripe sugarcane as received at the sugar mill is fed into a size reduction system where it is processed into sizes suitable for maximum juice extraction in a mill or a diffuser. This is done by the use of rotating knives, hammer mills or shredders. Usually, the washing of sugarcane stalk precedes the size reduction process; however, this is not the case in South African sugar mills (S. Davis, SMRI, personal communication). After size reduction, the processed sugarcane is then fed into a diffuser or a crushing mill for juice extraction where juice is separated from bagasse (fibre).

A mill system consists of multiple units of three-roller combination through which crushed cane or bagasse pass successively. The rollers are arranged in a triangular formation so that the cane fibre is crushed twice in each mill. To leach out as much sugar as possible from the cane fibre/bagasse, water or thin juice is sprayed on the blanket of bagasse as it emerges from each mill. This process is known as imbibitions. Two types of imbibition processes exist; simple and compound imbibitions (Hugot, 1986). Simple imbibition is where water is added to the bagasse after each mill unit. In compound imbibition, dilute juice (mostly water) obtained from the last mill unit or the last two or three mill units is returned to the mill unit that precedes it. More than 95% of sugar in the cane goes into juice in a best milling practice (Chen and Chou, 1993).

In a diffuser system, raw juice is extracted from the cane by the process of lixiviation using hot water and recirculation of the juice extracted for imbibitions (Hugot, 1986). It is the juice extraction system that is employed by most sugar mills in South Africa although three mills are still running milling tandems for juice extraction (S. Davis, SMRI, personal communication). Its extraction efficiency is 2-3% greater than that of a milling system and it also has lower maintenance cost (Modesto et al., 2009).

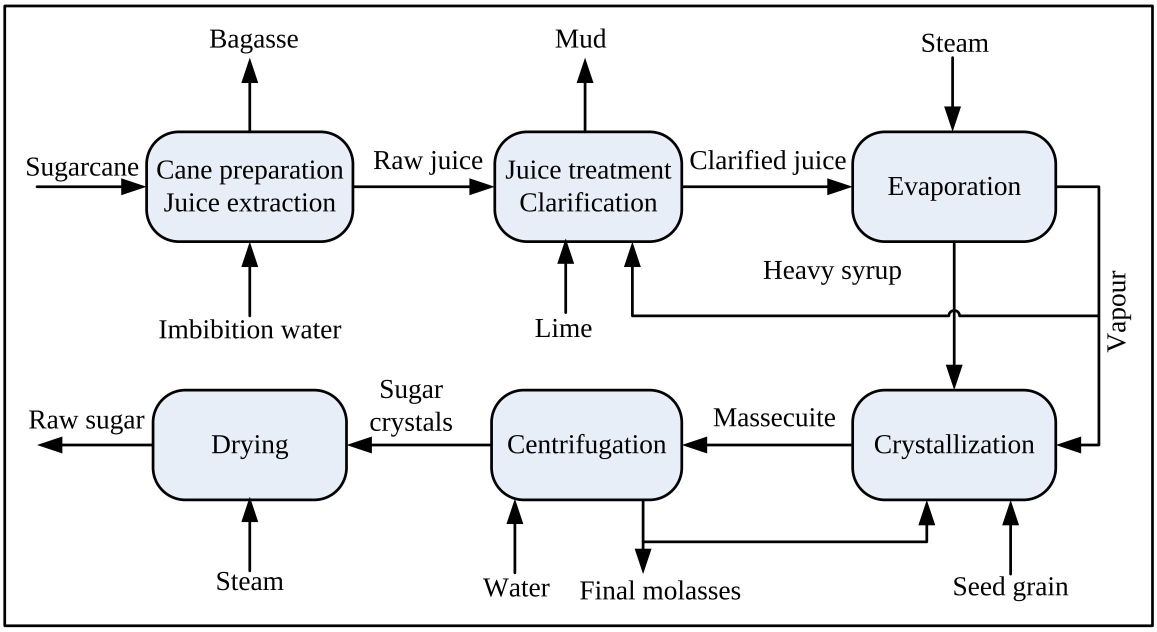

Figure 2: Schematic Representation of Raw Sugar Milling Process

2.2.2 Juice Treatment and Clarification

Raw juice obtained from sugarcane contains impurities such as fine particles of fibre, dirt, mineral salts and acids, besides sugar and water. These impurities have to be removed to enhance the purity of the final sugar obtained in the process. The purpose of the juice treatment and clarification step is to remove these impurities using lime and heat as clarification agents. In general, raw juice is first heated to raise its temperature (about 70^{o}C) followed by lime addition and then heated again to a much higher temperature (about 100^o – 105^{o}C) (Chen and Chou, 1993; Rein, 2007). However, the system of hot liming is used in South Africa, where raw juice is heated directly to 105^{o}C under pressure before lime addition. After this, the raw juice is flashed to eliminate air bubbles and then sent to a clarifier where it is separated into two streams: clarified juice and mud (contains mainly impurities and some proportion of sugar). The addition of lime neutralizes the acidity of the raw juice resulting in the formation of insoluble lime salt (calcium phosphate). This insoluble salt drags some other impurities during settlement, therefore enhancing the purity of the clarified juice (Mantelatto, 2005). Heating coagulates albumin and some waxes, gums, fats and the precipitates formed entrap suspended fine particles (Chen and Chou, 1993).

To ensure maximum sugar recovery, the mud is sent to a vacuum filter and the filtrate obtained is recycled to the process and mixed with the raw juice before lime addition. The filter cake is sent to the fields and used as fertilizer. Water is usually added in the filtration process to increase the filtration efficiency.

2.2.3 Evaporation

Clarified juice (containing about 15 wt% diluted solids) is concentrated to heavy syrup in a multiple–effect evaporator (five-effect evaporator in most industries). Exhaust steam from a cogeneration plant is used to provide the required thermal energy in the first evaporator effect. Vapour generated in this effect is used to provide the thermal energy required in the subsequent evaporator effects. Vacuum is imposed on the last evaporator effect and this makes the system to work in the order of decreasing pressures, and decreasing temperatures. To avoid sucrose loss and coloration, a maximum temperature of about 115^{o}C is set for the first evaporator effect (Baloh and Wittner, 1995). Hugot, (1986) also suggest a minimum pressure of 0.16 bar in the last effect. A fraction of vapour may be extracted from each effect and used as thermal energy source for other processes such as juice heating and sugar boiling.

2.2.4 Sugar Boiling/Crystallization

Heavy syrup containing about 65% sugar is concentrated in vacuum pans by boiling. Water is evaporated from the syrup until the syrup is saturated with sugar. Seed grain is added to the pan to serve as nuclei for the sugar crystals and more syrup added while boiling is continued. Crystal growth continues in crystallizers until the required crystal size is reached. This results in a dense mass known as massecuite (crystals and syrup mixture). Massecuite formed is fed to a centrifugal separation step where sugar crystals are separated from the syrup.

2.2.5 Centrifugation

In this step, raw sugar crystals are separated from syrup using a centrifuge which is basically a perforated basket revolving at high speed in a casing. Dark syrup otherwise known as mother liquor or molasses passes through the perforated lining of the centrifuge while the crystals are retained in the perforated lining. Water is added to wash the sugar crystals. Mother liquor is repeatedly boiled and centrifuged again until almost all available sugar crystals have been removed. In South Africa, three stages of crystallization and subsequent centrifugation are used to recover sugar. These are A, B and C crystallization/centrifugation stages, which produce A, B and C sugar and molasses, respectively. A molasses is fed to the B stage while B molasses is also fed to the C stage from which the final liquor (C molasses or final molasses) from the final boiling and centrifuging step is obtained.

2.2.6 Drying

Moisture content of sugar is reduced in a rotary sugar drier which consumes exhaust steam. Temperature control is very essential in this step.

2.2.7 Refining

Raw sugar contains some level of impurity and so have a brown colour. The purpose of the refining process is to remove virtually all the impurities and colour to form white sugar containing about 99.9% pure sucrose (Rein, 2007; Dias et al., 2009). The sugar refining process (see Figure 3) is similar to the raw sugar production process in terms of certain unit operations/processes such as evaporation, clarification, crystallisation, centrifugation, etc.

In South Africa, the sugar refining process starts with the melting/dissolving of raw sugar (see Figure 3) resulting in the production of raw liquor which is then clarified and filtered through a primary clarification and filtration process to remove colour and turbidity. Colour removal is enhanced through the use of either carbonatation and sulphitation or carbonatation and ion exchange processes or the addition of phosphoric acid (Hugot, 1986; Chen and Chou, 1993). The latter process is employed by only one sugar refinery in South Africa (S. Davis, SMRI, Personal communication). Lime is added to help maintain a proper pH during the clarification process (Hugot, 1986). After primary clarification and filtration, clarified liquor undergoes a secondary decolourisation step. In this step the clarified juice is passed through deep bed filters and then through ion-exchange resins for further colour removal, after which available ash is removed by softening resins. This results in decolourised liquor known as fine liquor. Fine liquor is then taken through the process of evaporation followed by crystallization and centrifugation to recover pure white sugar crystals, which are then dried and stored.

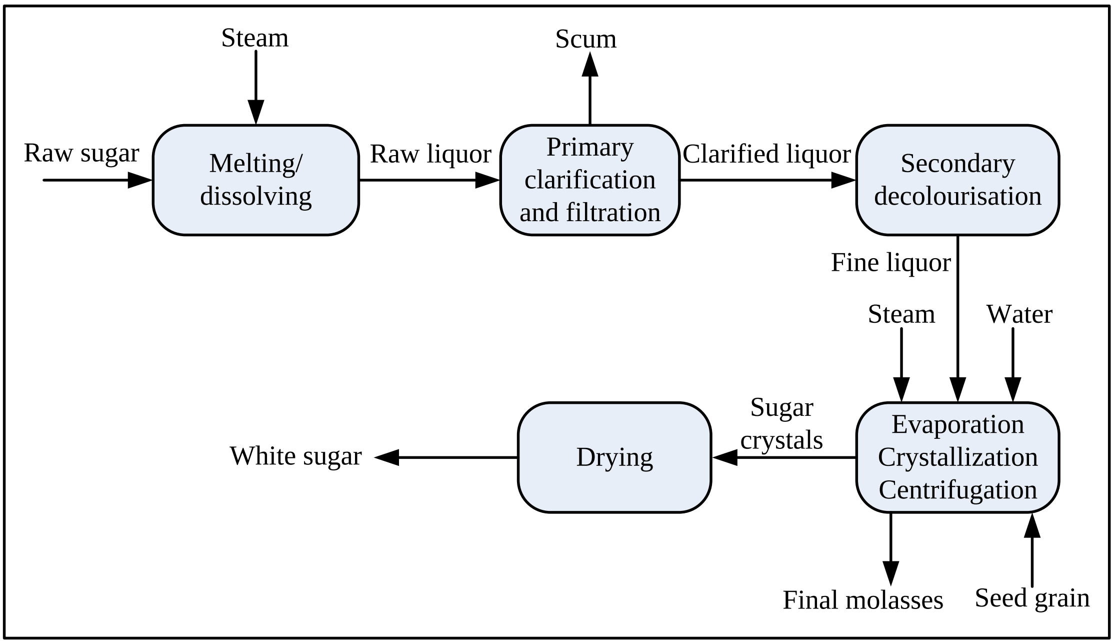

Figure 3: Block Flow Diagram of Sugar Refining Process

In countries where the purity of the raw sugar is very low, the refining process begins with a unit process called affination where clean raw sugar (affined sugar) is produced and then further refined to obtain pure white sugar. In affination, raw sugar is first mingled with heavy syrup which serves the purpose of softening the external layer of dried syrup on the raw sugar crystals. Crystals are then sent to affination centrifugals where softened layer is removed leaving clean raw sugar (affined sugar).

2.3 Process Energy Integration in the Sugar Industry

Almost all sugar cane processing plants nowadays have been designed to be energy self-sufficient (both thermal and electrical) with sugar as the main primary product. Bagasse, a by-product generated after juice extraction, is burned in bagasse–fired cogeneration systems to generate all the steam and electricity required to run the process. Due to the high levels of inefficiencies of most of these cogeneration systems and the lack of thermal integration of most sugar milling plants, little or no surplus bagasse is left (Ensinas et al., 2007).

With recent trends towards diversification in the sugar industry, several factories are manufacturing in addition to sugar, other by-products such as excess electricity, fuel grade alcohol, high-valued chemicals like furfural, etc. Each of these products requires a certain amount of energy or bagasse to manufacture. This implies that energy efficiency, both in the conversion of bagasse to useful energy and the use of this energy within the factory, can be of much benefit. Savings in energy would result in bagasse surplus. This could be used to produce by-products and this in turn will increase the profit margins of sugar mills. According to Botha and Blottnitz (2006), about 50% of the bagasse generated in a sugar mill can be saved if efficient energy integration measures are implemented within the mill. They based their argument on a hypothetical plant operating at 28% steam on cane.

Process energy integration is a useful tool for improving the recovery of energy from industrial processes. Several works have been done with regards to the application of energy integration in the sugar industry. The pinch method of analysis developed by Linnhoff (1982) and the method of applied exergetic analysis are the two methods that have been widely used in most of these works. Christodoulou (1992) evaluated the energy performance of a beet sugar factory through pinch technology. The study suggested the thermal integration of the process and the use of a six effects falling film evaporator. Tekin et al. (2001) assessed the effect of system operating parameters on the loss of useful energy/exergy in a beet sugar factory using structural bond coefficients. An increase in boiler efficiency and a reduction in the temperature of the exhaust steam were identified as critical measures to reduce energy loss.

Other works have focussed on cane sugar factories. Ogden et al. (1990) performed a thermal integration analysis of a raw cane sugar factory in Florida (USA), aiming at ways of economizing bagasse use and ensuring surplus bagasse for conversion into other useful products within the plant. The use of falling-film evaporators and continuous vacuum pans for sugar boiling were proposed to reduce process steam demand. Also improvement in the efficiency of the cogeneration system was proposed as a way of increasing electricity production. Mbohwa (2003) assessed energy integration measures adapted by cane sugar factories in Mauritius and Reunion. The study identified the use of high pressure steam generation systems and the use of vapour bleeds for heating purposes in low energy demanding processes as a means of minimising energy use and maximising its recovery. Ram and Banerjee (2003) used exergy and pinch analysis to evaluate two evaporation system designs for a sugar factory in India processing 5000 tons of cane per day. A modified evaporator design taking into account the reduction in evaporator surface area and the amount of steam consumed was proposed. It was concluded that modification of the existing quadruple effect with a modified quintuple effect will result in 9% and 48% reductions in steam consumption and exergy loss respectively.

Options for reducing the demand for thermal energy in sugarcane industries are presented by Rein (2007). These include:

- Maximum evaporation in multiple effect evaporators. In this way the concentration of solids in the syrup going to the pans is increased. This reduces the quantity of steam required for sugar boiling since much of the water would have been removed already. Quantity of vapour for heating auxiliary processes is also increased.

- Increase in the number of evaporator effects.
- The use of first, second and even third evaporator vapours in the pans.
- Minimum use of water for sugar washing.
- Using condensate to initially heat raw juice. This cools down condensate for use as imbibitions water in the mills.
- Use of vapours for juice heating.
- Increasing the temperature of bleed vapour so as to gain more from bleeding vapour.

Some recent research also points towards the use of reverse osmosis as a means of reducing the consumption of energy in the sugar industry. Reverse osmosis is a separation technique employing the use of membranes and operates without a change in phase, hence consuming a low amount of energy. Madaeni and Zereshki (2008, 2010) investigated the effect of using a two-stage reverse osmosis process as a pre-concentration step to partially separate water from thin juice prior to final concentration in the evaporation unit on the process energy demand. The result shows a considerable reduction in process energy for the use of reverse osmosis as compared to the conventional process which uses only evaporation for thin juice concentration.

2.4 Biomass

Plant biomass refers to any renewable source of energy produced from living creatures that stores energy by utilising the solar energy of the sun through photosynthesis. Biomass sources include waste materials generated from agricultural production and agro-processing activities, organic waste and crop residues. Others are forest products such as wood, sawdust, shrubs and tree bark. Energy crops such as herbaceous woody crops, sugar bearing crops like sugarcane and starch crops are also biomass sources.

Lignocellulosic biomass consist of three main components; cellulose, hemicellulose and lignin (Howard et al., 2003; Tsai et al., 2006; Dias et al., 2009; Saxena et al, 2009; Venderbosch and Prins, 2010; Sluiter et al., 2010; Basu, 2010). It also contains a small percentage of extraneous substances, predominantly organic extractives and inorganic minerals (Mohan et al., 2006).

Cellulose is a high molecular weight (about 10^6 amu or greater) polymer of D-glucose units linked by β-1,4 glucosydic bonds. It is an insoluble polymer and has a combination of both crystalline and amorphous structures (Hendriks and Zeeman, 2009; Goyal et al., 2008; Mohan et al., 2006). The molecules are held together by intramolecular and intrastrand hydrogen bonds, which is the reason why cellulose is insoluble in most solvents (Rowell, 1984). Together with hemicelluloses, they form the carbohydrate fraction of wood biomass known as polysaccharides.

Hemicellulose, also called polyose, is a polysaccharide polymer consisting of a mixture of sugars like glucose, mannose, xylose, galactose, arabinose, methylglucuronic acids and galacturonic acids. It is mostly located in the cell wall of plants where it bonds to lignin through covalent bonds and to cellulose through hydrogen bonds. Unlike cellulose, hemicellulose only has an amorphous structure and its average molecular weight is less than that of cellulose (Goyal et al., 2008; Mohan et al., 2006).

Lignin is a highly branched, high molecular weight, mononuclear, aromatic polymer, predominantly located in the cell walls of most plant biomass. It is a cross-linked amorphous resin with no unique structure and serves as a binder for the cellulosic fraction of biomass. It also provides a barrier against microbial or fungal attack on cellulosic components (Mohan et al., 2006). Lignin also provides structural support to plants (Hendriks and Zeeman, 2009). Its building blocks consist of phenylpropane units which are held together by ether and carbon-to-carbon bonds.

A typical woody biomass is made up of 65%-75% carbohydrates and 18-35% lignin. The extraneous matter (extractives and inorganic minerals) forms about 4-10% (Rowell, 1984).

2.4.1 Sugar Mill Biomass

Sugarcane bagasse and SCAR (sugarcane agricultural residues; consisting of trash, leaves and tops) are the main biomass produced in the sugar industry. It is estimated that 5.69 tons residue (wet basis) are generated per ton raw sugar produced, with SCAR and bagasse accounting for 42% and 45%, respectively (Font, 2000). Garcia-Perez et al. (2002) estimate that about 270kg (about 297kg for South Africa in 2010/2011 milling season (Smith et al., 2011)) of bagasse (50% moisture) is produced per ton of cane milled. Bagasse is one of the world's largest biomass sources and in South Africa, approximately 6 million tons of bagasse are produced annually (Leibbrandt, 2010; Hugo, 2010). Analysis on sugarcane bagasse by Garcia-Perez et al. (2002) shows that bagasse has the following composition: 35-50% cellulose, 20-30% hemicellulose, 20-27% lignin and 8-12% extractive and ash.

Though SCAR has a significant energy value comparable to bagasse, it is normally burnt in the fields before cane harvesting, while bagasse is utilised as combustion fuel in boilers at the mill to generate steam and electricity to run the mill.

2.5 Biomass to Energy conversion processes

The energy content of biomass can be harnessed through the use of one of these main processing routes: (i) Biochemical processes – anaerobic digestion and hydrolysis-fermentation – and (ii) Thermochemical processes. The biological processes are not discussed in this work as this work is more focussed on the use of thermochemical processes specifically, combustion and pyrolysis for the conversion of sugar mill biomass into energy and energy products.

2.5.1 Thermochemical Processes

Several thermochemical processes exist that can be used to convert biomass into useful energy. Some of these produce products that require immediate conversion while others produce products that can be stored for future conversion into energy. Thermochemical processes include gasification, combustion, liquefaction, hydrogenation, torrefaction and pyrolysis (Goyal et al., 2006; Bridgwater, 2011; Bridgwater, 2004). This work is aimed at the production of heat and electricity as well as energy dense products from sugar mill biomass, hence the focus is on combustion and pyrolysis, which are discussed in more detail, however a brief description of the other processes is also given.

Gasification

In gasification, biomass is converted into a mixture of combustible gases that can be combusted directly to produce heat for steam generation or can be used to run a gas turbine for the generation of electricity (Bridgwater et al., 2002). The process takes place under conditions of limited oxygen supply (partial oxidation) and high temperatures around 800-900^oC (Goyal et al., 2006). The gas mixture consists of carbon dioxide, hydrogen, carbon monoxide and methane. The gas products can be further processed to produce transport fuel through Fischer-Tropsch synthesis.

Though gasification can be used to produced heat and electricity, the technology is more advanced and requires more capital input (Bridgwater, et al, 2002).

Liquefaction

Here, biomass is converted into liquid products at low temperature and high pressure using hydrogen and in the presence of catalyst (Demirbas, 2000; 2001).

Hydrogenation

In this process, biomass is first gasified to produce syngas (CO and H_2 mixture) which is then subsequently converted to methane (Goyal et al., 2006).

2.5.1.1 Combustion

In combustion, biomass is fully oxidised at high temperature. Usually, an excess of oxygen (10-50% above stoichiometric value) is required to ensure that the combustion process is complete (Mbohwa, 2003). It is a well-established technology widely used for the provision of heat and electricity (Basu, 2010). The main product of combustion is heat, which cannot be stored and so must be readily used for heating purpose or electricity generation.

The simultaneous generation of heat and electricity through combustion is known as combined heat and power production (CHP). This process also referred to as cogeneration, has higher efficiencies than conventional power plants due to the use of waste heat as heat source in other processes (Magnusson, 2006; Al-Azri, 2008). The generation of heat and power can be done through the use of a steam turbine cycle, a gas turbine cycle or a combination of both (Bridgwater et al., 2002). In the steam turbine cycle, combustion fuel is combusted to produce hot gases, which is then used to generate steam. The steam is let down through a turbine that is connected to an electric generator to generate electricity. The residual steam from the turbine is then used for heating purposes. In the gas turbine, combustion gas is directly expanded in a turbine system for power production and the residual gas used for heating. A combination of both the steam and gas turbine cycles constitutes the combined cycle. In the combined cycle, high temperature combustion gas is first expanded in a gas turbine and then the exhaust gas used to generate steam which is then also expanded in a steam turbine and the exhaust steam used for heating duties. Among these options, the steam turbine CHP system is the most well established and widely used technology (Bridgwater et al., 2002). Its advantages include the ability to use different types of fuels, high heat to power ratio; 2-10 kW_{th} /kW_e compared to 0.5-2 kW_{th}/kW_e for gas turbine, and a year round plant availability (Savola and Fogelholm, 2006; EDUCOGEN and INESTENE, 2001; Mani et al., 2010). Due to these reasons, the steam turbine based cogeneration system has found wide application in most sugar mills around the world for the conversion of bagasse into steam and electricity to run the milling operation. A steam cycle based CHP plant is thus discussed briefly below.

2.5.1.1.1 CHP Plant Based on Steam Cycle

This type of CHP plant is made up of a steam boiler, steam turbine and a power generator (Magnusson, 2006). High pressure steam (HP steam) is first produced in the boiler and then let down through a steam turbine. In the turbine, the steam expands and is used to run a generator for electricity production (Mbohwa, 2003; Pippo et al., 2009). Exhaust steam exiting from the turbine leaves at a low pressure and temperature due to the expansion process and is commonly referred to as low pressure steam (LP steam). In an ideal CHP plant (Figure 4), this steam is condensed and recycled back to the boiler. However in the case of the sugar mill, this exhaust steam is used in the mill to run processes like juice evaporation, sugar boiling and other heat demanding tasks.

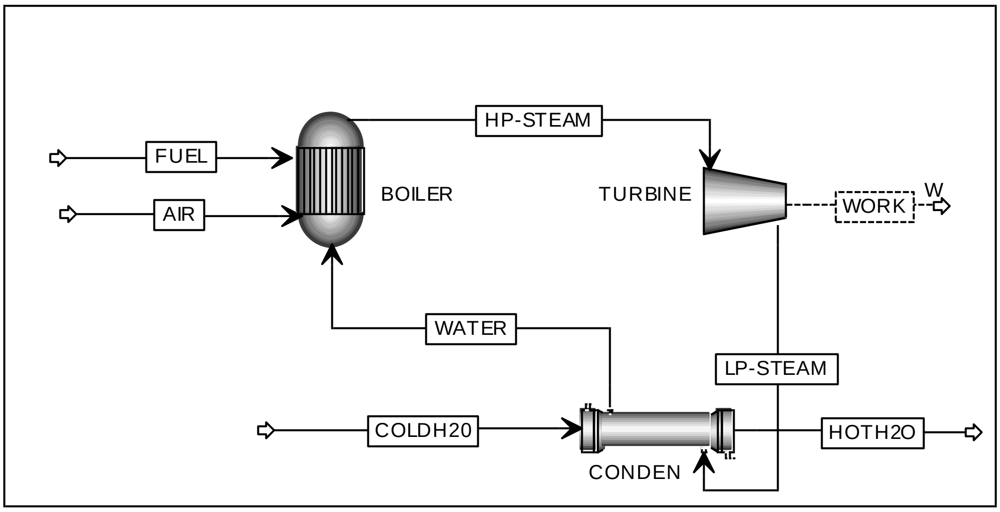

Figure 4: PFD of Ideal Steam-Turbine CHP Plant (redrawn from Magnusson, 2006)

Steam boiler

The steam boiler basically consist of a combustion chamber and a vessel containing water where steam is raised from the water through heat exchange between the water and hot gases from the combustion chamber. A fuel–air mixture is fed to the combustion chamber. To ensure complete fuel combustion, an excess of air (10-50%) above the stoichiometric amount is provided. Flue gas generated is then used to produce superheated steam from boiler feed-water, which then goes to the turbine cycle. To maintain a constant pressure within the boiler, the amount of steam leaving the boiler must be equal to that produced from the heat supplied to the boiler feedwater (Magnusson, 2006). The pressure drops if the amount of steam exiting exceeds the amount produced and it increases if the amount produced is greater than that which is exiting. To improve the efficiency of the boiler, both the feed water and the combustion air must be preheated (EDUCOGEN and INESTENE, 2001). Air preheating is essential especially when using a wet fuel.

The steam turbine

To generate electrical power, high pressure steam produced from the steam boiler is expanded in a steam turbine and used to run an electricity generator. The efficiency of the turbine and hence the quantity of power produced depends on the inlet conditions of the turbine. Higher steam pressure results in higher conversion efficiency of thermal energy to electrical power. However, this requires greater boiler capacity and operating cost, a trade-off thus needs to be made between power production and cost (Pippo et al., 2009).

Two kinds of steam turbines are commercially available; back-pressure turbines and condensing turbines. In a back-pressure turbine, exhaust steam from the turbine have pressure that is above atmospheric pressure and has higher energy content in terms of heat. Its electrical output is usually less. To increase the electrical output, a back-pressure turbine can be connected to a condensing turbine. In this arrangement, high pressure steam is first expanded in the back-pressure turbine and then let down through a condensing turbine. Some quantity of steam however may be extracted from the back-pressure exit before the condensing turbine depending on the heating requirement of the end user (Quevauvilliers, 2001).

The condensing turbine produces an exhaust steam with pressure lower than atmospheric pressure and requires a condenser. Steam exiting from the turbine is passed on directly to a condenser and the condenser temperature determines the pressure of the steam at the turbine exit. Condensing turbines are designed to maximise electrical output (Quevauvilliers, 2001; Pippo et al., 2009). Just like the back-pressure turbine, steam can also be extracted from a condensing turbine for heating purposes; however, this decreases the power output.

Generators

The purpose of the generator is to produce electrical energy from the mechanical energy of a turbine. The type used in cogeneration plants is the synchronous generator. This is because this type of generator can operate as a stand-alone plant without been connected to other generators or the grid. Another type, asynchronous generators can only function alongside other generators.

2.5.1.2 Pyrolysis

Pyrolysis is a thermochemical process used to convert biomass into higher value products. It takes place in the absence of oxygen and normally in an inert atmosphere (Lu et al., 2009; Bridgwater, 2011) and this differentiates it from other thermochemical processes such as gasification and combustion, which require partial and complete oxidation, respectively (Basu, 2010; Goyal et al., 2008). The products of pyrolysis are a mixture of condensable gases (this gases condenses to form bio-oil), non-condensable gases and char, which is an energy rich solid product (Venderbosch and Prins, 2010; Bridgwater et al., 2003). There are different modes of pyrolysis; slow, fast and vacuum pyrolysis. The difference between these technologies is the temperature and rate at which heat is transported to the biomass and the rate at which vapour is removed. Heating rates and temperature determines the extent to which the various components of biomass decompose while the short vapour residence time and rapid cooling prevents the secondary cracking of vapours and determines the yield of pyrolysis products (Venderbosch and Prins, 2010; Basu, 2010; Bridgwater, 2011; Bridgwater et al., 1999).

Biomass is composed mainly of cellulose, hemicellulose, lignin and some small amount of other organics. Each of these components decomposes at different rate and mechanism when biomass is exposed to the conditions of pyrolysis. Cellulose and hemicellulose degrade very quickly over narrow temperature ranges, while lignin decomposes over a wider range of temperature (Bridgwater et al., 1999). The original molecular bonds and chemical structure of biomass break down during pyrolysis, producing mostly oxygenated compounds of varying molecular weights. A greater proportion of these compounds condense to form bio-oil, while those with lower molecular weights such as CO_2 remain as permanent gases at ambient temperature (Ringer et al., 2006).

Biomass pyrolysis is of growing interest and has for the past three decades been a subject of intensive research (Bridgwater et al., 1999; Meier et al., 1999; Czernik et al., 2004; Ringer et al., 2006; Garcia-Perez et al., 2010; Venderbosch and Prins, 2010; Bridgwater, 2011). Unlike other thermal conversion processes, which require the immediate use of their products, pyrolysis products can readily be stored and used when the need arises. They can also be transported to centralised refineries for upgrade into transportable fuels and synthesis of valuable chemicals or to power stations for the generation of power. Pyrolysis therefore offers logistical and practical advantages over other thermochemical processes. About 70% of biomass feed by mass ends up in bio-crude/bio-oil during biomass pyrolysis (Garcia-Perez et al., 2010; Oasmaa and Peacocke, 2010; Venderbosch and Prins, 2010). The bio-oil, which has a dark brown colour, also has very similar elemental composition as the biomass from which it is produced. Bio-oil contains 10-25 mass % water and between 75-95 mass % organic compounds (Briens et al., 2008; Bridgewater, 2011). It also contains small amounts of ash and solid carbon/char due to the inefficient separation of char from the pyrolytic vapours by cyclones often used for this purpose.

The distribution of product yields depends on the mode of pyrolysis that is applied to the biomass feed. Fast pyrolysis maximises liquid product yield, which is favoured by the high heating rates and short vapour residence time employed. Up to about 80% (wet mass basis) liquid yield is achievable on dry feed (Bridgwater et al., 1999). Bridgwater et al. (2001) obtained on weight basis a product distribution of 75%, 12% and 13% of bio-oil, char and gas, respectively, from fast pyrolysis of wood. Le’de’ et al. (2007), obtained similar yields of 74%, 10% and 16% for bio-oil, gas and char, respectively from fast pyrolysis of wood sawdust in a cyclone reactor. Piskorz et al. (1998), obtained bio-oil yield of 60% (energy basis) from fast pyrolysis of sugarcane bagasse. Slow pyrolysis favours the formation of solid products due to lower heating rates and long residence time. A typical product distribution between bio-oil, char and gas on weight basis as obtained by Bridgwater et al. (2001) for slow pyrolysis of wood is 30%, 35% and 35%, respectively. Vacuum pyrolysis also favours the formation of solids and gases as opposed to liquid products; however the distribution is more evenly spread and the bio-oil is of a higher quality. Table 2 presents a typical pyrolysis product distribution for biomass under different pyrolysis modes.

Table 2: Pyrolysis Product Distribution for Biomass under different Pyrolysis Modes

Mode of Pyrolysis	**Comment**	**Product distribution, wt %**		
		Liquids	**Gas**	**Char**
Fast	500^{o}C, high heating rate, short vapour residence time (less than 2s)	60 - 75	10 - 20	10 - 20
Vacuum	450^{o}C, low heating rate, long – medium vapour residence time(few min)	35 - 45	25 - 35	20 - 30
Slow	500^{o}C, low heating rate, long- long vapour residence time (5min – 30min)	30 - 45	30 - 45	25 - 35

Source: Bridgewater et al., 2003

The different modes of biomass pyrolysis are described below. Slow and vacuum pyrolysis modes are briefly described, while more emphasis is placed on fast pyrolysis as it pertains to this work.

2.5.1.2.1 Slow Pyrolysis

Slow pyrolysis, also called carbonization, is a well-known, conventional pyrolysis process, application of which dates back several thousands of years (Bridgwater, 2011, Bridgwater, 2004). It was applied in some form for the traditional production of charcoal (Bridgwater, 2011; Venderbosch and Prins, 2010). Low heating rates and hence slow heat transfer rates to biomass and long vapour residence time (5 – 30 min) characterise slow pyrolysis (Bridgwater et al., 2003). It takes place at temperatures ranging between 400-500^{o}C in an inert atmosphere. The feeding of nitrogen gas through the pyrolysis reactor creates the inert atmosphere and also controls the vapour residence time, due to the slow rate at which the gas is fed. The slow rate at which heat is imparted to the feed and the long vapour residence time causes secondary cracking of product vapours, which reduces the yield of organic liquid. Hence slow pyrolysis favours the formation of solid char, which is considered as the main product. Two main technologies are available, i.e. kiln technology and slow pyrolysis retort. Kiln technology produces charcoal as the sole product, whilst the retort technology produces charcoal together with some small quantities of bio-oil and gas. The bio-oil and the gas are normally combusted to generate heat for the process. Both technologies require the use of high density particles as they tend to favour an increase in char yield (Erlich et al., 2006).

2.5.1.2.2 Vacuum Pyrolysis

In vacuum pyrolysis, biomass feed is thermochemically converted into bio-oil, gas and char under vacuum (Goyal et al., 2008). The typical temperature range is around 350°C - 520°C and the vacuum pressure employed is in the range of 10 – 20 kPa (Rabe, 2005). Vacuum pyrolysis differs from fast and slow pyrolysis processes as it requires no carrier gas and operates under vacuum, while the other pyrolysis processes operates at normal atmospheric conditions and makes use of a carrier gas. It can be regarded as both a slow pyrolysis process and a fast pyrolysis process. Slow because it is characterised by slow heating and heat transfer rates (Bridgwater, 2011; Bridgwater et al., 1999). Fast because the vacuum ensures the rapid removal of pyrolysis products from the reactor (Goyal et al., 2008) thus giving it a shorter vapour residence time similar to that of fast pyrolysis (Bridgwater, 2011), which results in the reduction of secondary reactions of volatile vapours. Because of the slow heating rate, the yield of liquid products (dry feed basis) from vacuum pyrolysis is in the range of 35-50 wt% (Bridgwater, 2011; Basu, 2010). This is much lower when compared to 70-75 wt% obtainable from fast pyrolysis using a fluidized bed reactor (Bridgwater, 2011).

The advantages of vacuum pyrolysis include but are not limited to the following (Ringer et al., 2006; Bridgwater et al., 1999; Bridgwater, 2011):

i. the production of bio-oil with less char particles.
ii. the ability to use larger feed particle sizes than fluidized bed reactors mostly applied in fast pyrolysis.
iii. the elimination of the use of a carrier gas to remove vapours rapidly from the pyrolysis reactor thus reducing the formation of aerosols (Ringer et al., 2006).

A major disadvantage is the high operating cost associated with the vacuum system (Bridgwater, 2011). Also the low liquid yield makes it unsuitable for liquid fuel applications (Ringer et al., 2006).

2.5.1.2.3 Fast Pyrolysis

In fast pyrolysis biomass is heated rapidly to high temperature in the absence of oxygen, resulting in the conversion/decomposition of the biomass into a mixture of condensable and non-condensable gases and solid char. Cooling of the gases produces a liquid product (bio-oil), which condenses out of the condensable gas and has a heating value of 14-18 MJ/kg (LHV), about half that of fossil based heavy fuel oils (41-43 MJ/kg, LHV). Fast pyrolysis is controlled to maximise liquid product yield. The keys to obtaining high yields of bio-oil in a fast pyrolysis process include a very well controlled pyrolysis reaction temperature (~500°C), with short vapour residence times (<2s), very high heating rates (10^3-10^5 °C/s) and fast heat transfer rates (achievable through the use of smaller biomass particle feed sizes), and rapid quenching of the pyrolysis vapours (Bridgwater et al., 1999).

Bio-oil, the main fast pyrolysis product, is a miscible mixture of polar organics and water. Its moisture content ranges from 15 to 30 wt%, depending on the initial amount of water in the feedstock used and also on the pyrolysis conditions (Solantausta et al., 1993; Lu et al., 2009). Fast pyrolysis can be said to be a waste-free process. Char and gas which are by-products can be combusted to provide the high heat requirement of the process; hence no waste streams are produced, except flue gas (composed predominantly of CO_2 and water), from the combustion of by-products (Bridgwater, 2011).

Several reactors have been developed for fast pyrolysis application, however, fluidized bed reactors are the most widely used, because they are simple and easy to operate and scale-up and they have very good heat transfer rates to biomass (Bridgwater, 2011; Venderbosch and Prins, 2010). Figure 5 shows a fluidized bed fast pyrolysis process setup. In this process, biomass is first dried to about 10% or less moisture content using heat from the by-products, followed by size reduction (to < 2 mm) in order to increase the rate of heat transfer (Bridgwater, 2011; Bridgwater et al., 1999). After this the prepared biomass is fed into the pyrolysis reactor for pyrolysis under carefully controlled fast pyrolysis conditions. The heat for pyrolysis is supplied by the combustion of part of the by-products. The products of pyrolysis are then sent first through a cyclone to separate the solid char from the vapours and then the vapours rapidly quenched to recover the bio-oil from the non-condensable gas.

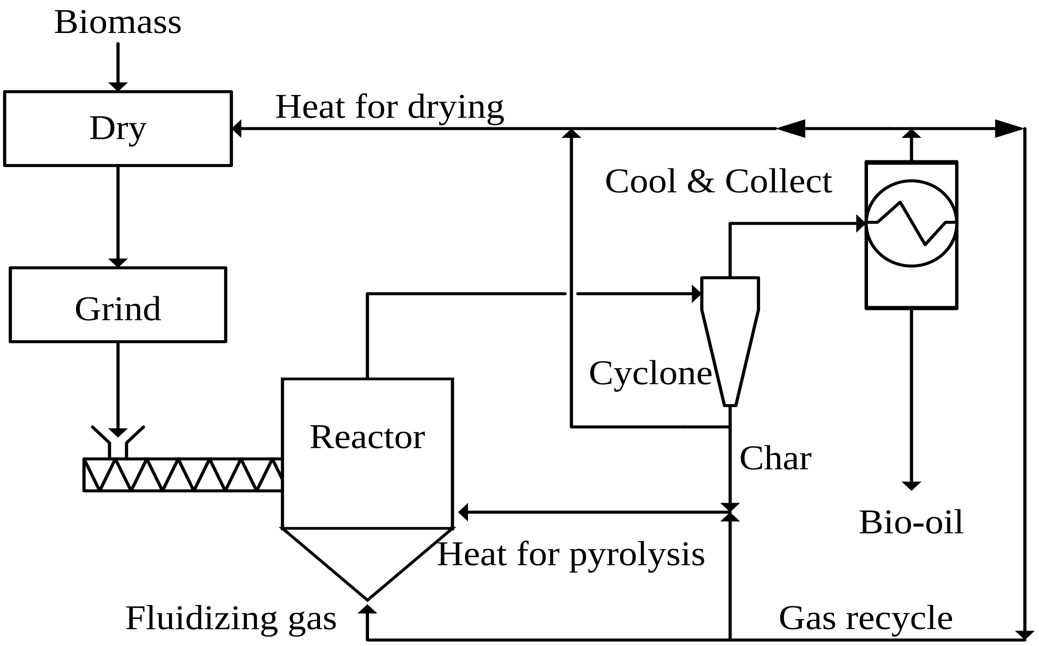

Figure 5: Fluidized bed fast pyrolysis process (redrawn from Bridgwater et al., 1999)

2.5.1.2.3.1 Fast Pyrolysis Process Characteristics and Technology Requirements

Even though conventional pyrolysis had been in existence for so many years, fast pyrolysis an advanced technology only emerged around the 1980's, and has for the past three decades seen significant research and development (Bridgwater et al., 1999) by researchers with regards to process operating parameters and technology. All these research efforts have aimed at improving the performance, reliability, and product yield, quality and consistency of fast pyrolysis. This section thus presents some of the process characteristics and technological requirements for achieving these goals.

Heating and Heat Transfer Rates

To obtain higher liquid yields, very high heating and heat transfer rates are required within a very short time followed by rapid cooling of products after their formation (Ringer et al., 2006; Lu et al., 2009; Bridgwater, 2011; Bridgwater et al., 1999). Heating rate of about 10^3-10^5 $^oC/s$ (Lu et al., 2009; Basu, 2010) and less than 2s residence time (Bridgwater, 2011; Bridgwater et al., 1999) are recommended. High heating and heat transfer rates favour the formation of vapours and hence the increase in liquid yields, while low heating rates favour char formation as it occurs in slow pyrolysis process.

The transfer of heat to biomass in a pyrolysis reactor is in two steps; first to the heat transfer medium of the reactor and then to the biomass (Bridgwater et al., 1999; Basu, 2010). The main heat transfer mechanisms during pyrolysis are convection and conduction, however, radiation also plays a critical role as heat is first transferred to the exterior surface of biomass by radiation and convection after which conduction and convection take over the transfer of heat to the interior portion of the biomass (Babu and Chaurasia, 2004). The magnitude of the contribution of the conduction and convection modes of heat transfer to the pyrolysis process depends on the type of reactor used. Each mechanism puts some limitations on the operation of the pyrolysis reactor (Bridgwater, 1998; Bridgwater et al., 1999), and hence a maximum contribution from each one or both must be made in order to maximise the yield of liquid products. For example, an entrained flow reactor uses convection as its primary mode of heat transfer, since heat is transferred from hot gas to the biomass particle. The ablative reactor maximises the use of the conduction. Heat transfer is mainly from the reactor walls to the biomass particles. Fluidised bed and circulation fluidised bed reactors employ significant contributions from both heat transfer modes (see Table 3), i.e. solid-solid transfer from good solid mixing and gas-solid transfer from hot fluidising gas.

Table 3: Reactor Types and Heat Transfer Modes (reproduced from Bridgwater et al., 1999)

Reactor type	**Heat transfer mode, %**		
	Conduction	***Convection***	***Radiation***
Ablative	95	4	1
Circulating fluidized bed	80	19	1
Fluidized bed	90	9	1
Entrained flow	4	95	1

Feedstock Preparation

Achieving high heat transfer rates to biomass, as required for fast pyrolysis, by applying high heating rates, is dependent on the nature of the biomass feedstock used, with regards to particle size and moisture content. Adequate preparation of feedstock is thus a necessity. The heat transfer modes pertaining to fast pyrolysis of biomass require that relatively small particles be introduced to the reaction vessel hence a limitation is imposed on particle size. Small particle size provides an increased surface area per unit volume of particle, hence ensuring that the required pyrolysis temperature is reached throughout the particle in a short time. Also the formation of char and its subsequent deposition at the surface of biomass particle during pyrolysis is another reason for requiring small particles. Char is insulating in nature (Ringer et al., 2006), and therefore hinder the rate at which heat is transferred into the inner portions of particles. Hence the need for smaller particle size to offset this effect. Biomass have very poor thermal conductivity (0.1 W/mK along the grain, ca. 0.05 W/mK cross grain) according to Bridgwater et al. (1999). This implies that to meet the requirements of rapid heating through gas-solid heat transfer, biomass particle must be as small as possible. Scott and Piskorz, (1984), investigated the effect of particle size on the yield of bio-oil and concluded that increasing particle size reduces the yield of bio-oil as there is a significant increase in secondary reactions within the particle.

Biomass particle size of about 2mm or less is appropriate for fast pyrolysis (Bridgwater et al., 2001), although some reactor types, such as the ablative reactor, can accept large size particles. Scott and Piskorz (1982) identified that product yields are not affected much when particle sizes are smaller than 2mm. The cost of energy required for size reduction must however be considered as this has a direct effect on the overall cost of production. Furthermore, sand /soil particles need to be removed from biomass before size reduction in order to reduce the overall ash content of biomass which might negatively affect the quality of the bio-oil as well as slagging in the reactor (Lu et al., 2009). This additional processing adds to the cost of biomass processing.

The moisture content of feed must also be considered during feedstock preparation as any moisture present only vaporises during pyrolysis and re-condense with the bio-oil, thereby affecting the quality of the oil. According to Ringer et al. (2006) a bone-dry biomass will produce bio-oil having a moisture content of about 12-15 wt% water. Moisture in a wet feed will thus be added to this initial amount, thereby increasing the overall moisture content of the bio-oil. Moisture in the feedstock acts as a heat sink inside the pyrolysis reactor (Ringer et al., 2006), since heat is required to evaporate any available water. Increasing moisture content of feedstock thus decreases the amount of heat inside the pyrolysis reactor that is available for biomass pyrolysis, and this in turn decreases the rate at which heat is transferred to the biomass itself. In other words, higher feed moisture content requires that more heat energy is supplied to the pyrolysis reactor, resulting in a negative effect on the overall energy balance of the process.

Drying of biomass feedstock is therefore necessary. Drying to 5-10 wt% moisture content is acceptable for fast pyrolysis (Bridgwater et al., 1999; Ringer et al., 2006; Lu et al., 2009).

Temperature

Temperature in pyrolysis can be understood in two terms; the reaction temperature and the temperature of the reactor. The principle of heat transfer requires that a temperature gradient be created; hence the reactor temperature must always be greater than the reaction temperature. Fast pyrolysis of biomass has been optimised for maximum yields of liquid products and the temperature range 500-520 ^{o}C has been determined to be the optimum for wood based biomass to achieve this goal (Bridgwater et al., 1999; Basu, 2010; Venderbosch and Prins, 2010). Experimental work by Hugo (2010) reports a temperature range of 495-510 ^{o}C as the optimum for obtaining maximum liquid product yields from fast pyrolysis of sugarcane bagasse. Temperature affects the total product yield of most wood based biomass during pyrolysis hence its control is of critical importance. The various components of biomass, as stated above, degrade at different temperature ranges, implying that only certain components do react at a particular pyrolysis temperature, thus affecting yield and product quality (Venderbosch and Prins, 2010). Lower temperature favours the formation of char and decreases the yield of liquid products because of low heating rates and incomplete pyrolysis (Onay, 2007). However, at high temperatures (>500^{o}C) a decrease in liquid yields may also occur when vapour residence time is long as a result of secondary cracking of organic vapours into non-condensable gases. The effect of temperature on the yield of fast pyrolysis products is presented in Figure 6 below.

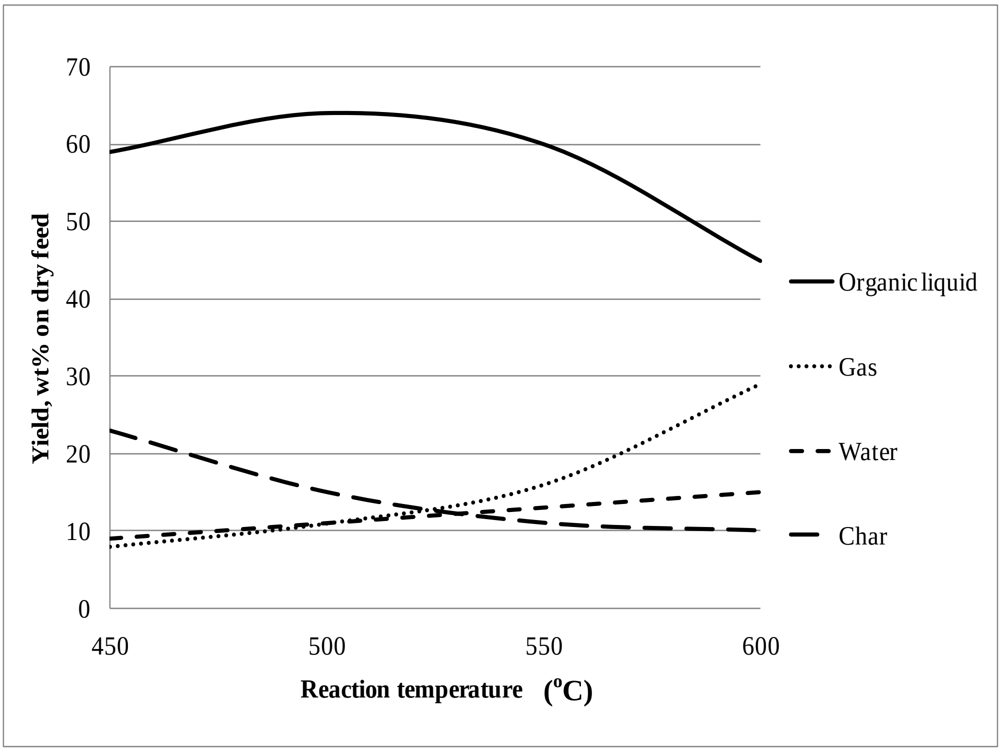

Figure 6: Effect of temperature on the yield of products from fast pyrolysis of wood (reproduced from Bridgwater et al., 1999)

Reactor Types

The choice of reactor for fast pyrolysis applications is of much importance as it determines to some extent the processing parameters (temperature, pressure, etc.) required. It also indirectly affects the processing cost of the biomass feedstock and the overall process operating cost, as different reactors require special feedstock particle sizes. Several reactors have been developed and investigated for fast pyrolysis application since inception of the technology, to determine which one provides the high heating rates, moderate temperatures and the short vapour residence times that are the necessary attributes for producing high liquid yields. These are grouped under the following:

- Fluidized bed
- Transported bed
- Circulating fluid bed
- Ablative (vortex and rotating blade)
- Rotating cone
- Vacuum

Of these reactor configurations, the fluidized bed has proven to be the most widely used because it can be scaled-up easily and it is easy to operate (Bridgewater, 1999; Bridgwater, 2011). It however requires very small particle sizes (<2mm) to obtain good yields of liquid product. This is a disadvantage when cost of feedstock preparation is taking into account. Also the use of a substantial amount of carrier gas for fluidization has an effect on liquid product recovery. The carrier gas contributes to aerosols formation during thermal quenching of the process stream and this in turn makes the liquid oil recovery more difficult (Ringer et al., 2006).

Vapour Residence Time and Secondary Cracking

Vapour residence time, defined as the time a molecule spends on average in the reaction environment of the reactor, has a substantial effect on the yield and quality of the organic liquid product (bio-oil). Research has shown that at longer vapour residence times, yields of organic liquids decreases while that of product water, char and gas increases. This is as a result of secondary cracking occurring within the hot reactor environment caused by the catalytic active nature of char. Organic vapours are converted to secondary char, water and non-condensable gases if allowed to stay for a long time in the reactor environment (Bridgwater et al., 1990). The rapid removal of products is therefore very important. The effect of residence time is also linked to temperature. Research work by both Diebold et al. (1987) and LideAn et al. (1988) investigated the relationship between the formation of primary liquids and secondary cracking, but both studies disregarded the variation of water formation with temperature and residence time. There is a decrease in the average molecular weight of liquid products at temperatures less than 400^{o}C as a result of secondary condensation reactions (Bridgwater et al., 1999) while Boroson et al. (1989) showed that the decrease was inversely related to temperature and residence time. Increasing the vapour residence time and temperature leads to secondary cracking of organic vapours and hence a reduction in the yields of organic liquids and specific products. Typically, vapour residence time of 2s or less is recommended for obtaining high organic liquid yields at the optimum pyrolysis temperature (500^{o}C) (Bridgwater et al., 1999; Ringer et al., 2006; Lu et al., 2010).

Bio-oil Recovery/Collection

In order to preserve the chemical composition of the pyrolysis vapour/bio-oil, and also prevent vapours from cracking further to permanent gases or polymerize to secondary char (which in turn affects the quality and yield of bio-oil), it is required that pyrolysis vapours be rapidly removed from the reactor system once formed and thermally cooled/quenched from the high reaction temperature. Pyrolysis vapours do create problems during cooling and condensation due to their nature. They exist as mist at low concentrations in an inert gas, and hence are very difficult to capture (Bridgwater et al., 1999). When cooled, pyrolysis vapours form aerosols (Ringer et al., 2006), which are very difficult to remove from the permanent gases due to their submicron sizes. This problem is even more significant when larger quantities of carrier gas are present during condensation. Venturi scrubbers have been effectively used to capture aerosols, although they pose a problem of high pressure drop (>10 kPa), which may not be available from the pyrolysis process (Ringer et al. (2006). Currently electrostatic precipitators are used for aerosol capturing but they are very expensive.

Temperature and the rate of cooling are very critical design considerations during liquid recovery. Slow cooling rates (using simple heat exchange) lead to liquid fractionation and eventual blockage of heat exchange equipment, resulting from the preferential collection of viscous lignin derived components (Bridgewater et al., 1999). Rapid cooling/quenching achieved through direct contacting of vapours with cooled liquids is effective, but temperature-dependent as the temperature determines the yield and quality of the bio-oil obtained. At lower temperature, there is an increase in yield because more moisture and volatile organic vapours are condensed. High temperature produces a highly viscous but a high calorific value bio-oil.

Char and Particulate Separation

Char, a co-product of fast pyrolysis, is catalytically active (Bridgwater et al., 1999). It acts as a catalyst to speed up most of the secondary reactions that occur during pyrolysis. The effective and swift separation of char from pyrolysis vapours is therefore very important to ensure high yields of liquid products. It is more desirable that char be separated from the vapour stream before cooling and condensing vapour into liquid (Ringer et al., 2006). Cyclones are normally used to separate char from vapours as they exit from the high temperature reaction vessel. However, cyclones are inefficient as they have a limitation with regards to particle size. Even a very well designed cyclone is ineffective on particles whose sizes are below 2-3 microns (Ringer et al., 2006; Lu et al., 2009; Garcia-Perez, 2010), implying that some fine char particles make their way through the cyclone and end up in the liquid products causing problems of instability to the bio-oil (Bridgwater, 2004; Oasmaa et al., 2003; Lu et al., 2009). Currently research is underway to find possible ways of effectively removing char from pyrolysis vapour as a means of improving on bio-oil quality. A technique using hot gas filtration *in lieu* of a cyclone to remove almost all char from pyrolysis vapour prior to condensation have been developed (Hoektra et al., 2009; Scahill et al., 1997; Diebold et al., 1994) . The technique makes use of a bag-house filter with small modifications to ensure minimum residence time of vapour within the filter. Although this technique produced a high quality, char-free bio-oil, there was a 10-15% loss in liquid yield due to vapour cracking caused by char accumulation on the surface of the filter. Other techniques that are still under investigation include in-bed filtration (Wang et al., 2004) and the use of rotary particle separators within the pyrolysis reactor (Bramer et al., 2004).

A summary of the process characteristics and technological requirements for fast pyrolysis as discussed above is given in Table 4 below.

Table 4: Summary of Process Characteristics and Technology Requirements for Fast Pyrolysis

Parameter	Comment
Feedstock preparation	
Feed drying	Essential to ≈10% (less moisture desirable)
Particle size	Smaller particles needed (<2mm), costly
Sand removal	Essential for fibrous biomass (might lead to increased ash content)
Reactor	
Reactor type	Many configurations exist (fluidized bed mostly used)
Heat transfer	Gas-solid and/or solid-solid
Heat supply	High heat transfer rate required
Heating rate	High heating rates needed; Limited by the thermal conductivity of biomass
Reaction temperature	500^{o}C maximises liquid yields
Product recovery	
Vapour residence time	Shorter residence times essential (<2s)
Secondary cracking	Reduces liquid yields; not desirable
Liquid collection	Difficult; rapid quenching and electrostatic precipitation seems best
Char separation	Difficult; hot gas filtration seems best
Ash separation	More difficult than char; hot gas filtration seems best.

2.5.1.2.3.2 Product Characteristics, Properties and Composition

Bio-oil or bio-crude is the single most abundant product of biomass fast pyrolysis. It is a free flowing dark brown coloured liquid, with a typical smoky odour and is generally homogeneous in appearance (Bridgwater, 2011; Venderbosch and Prins, 2010; Bridgwater et al., 1999). It is a complex mixture of compounds produced when cellulose, hemicellulose and lignin undergo depolymerisation and fragmentation reactions under the conditions of pyrolysis (Bridgwater et al., 1999; Bridgwater, 2004; Bridgwater, 2011; Lu et al., 2009). Chemically, it is composed mainly of water, some solid particles (value depending on the efficiency of the solid separation method used) and several organic compounds belonging to the following functional groups: acids, ketones, aldehydes, alcohols, phenols, ethers, esters, sugars, furans, alkenes, aromatics, nitrogen compounds and multifunctional oxygenate compounds (Lu et al., 2009, Milne et al., 1997; Diebold, 1997). The overall chemical composition of bio-oil depends on factors such as the type of biomass used in producing the oil, feedstock preparation (particle size and moisture content), conditions of pyrolysis as well as the vapour filtration and condensation mechanism employed in obtaining the oil (Venderbosch and Prins, 2010; Bridgwater, 2011). This implies that the composition of bio-oils produced from different biomass feedstocks and different reactor configurations differ greatly. On an elemental basis, bio-oil is very similar to the biomass feedstock from which it is produced in terms of composition (Bridgwater, 2011) as can be seen from Table 5.

Table 5: Typical Elemental Composition of Fast Pyrolysis Bio-oil compared to that of some Biomass Feedstocks

Component	**Composition, wt%**		
	Bio-oil[a]	Sugarcane bagasse[b]	Hard wood[c]
C	44-47	47.4	49.9
H	6-7	7.2	6.0
O	46-48	40.69	43.8
N	0-0.2	n.a	0.3

([a]Bridgwater et al., 1999; [b]Fernandez et al 2004: in Pippo et al., 2007; [c]Garcıa-Perez et al.,2002)

Generally the oxygen content of bio-oils varies in the range 35-60 wt% on wet basis (Lu et al., 2009). A range of 45-50 wt% is quoted by Oasmaa et al. (1999) and Bridgwater, (2004) while Ringer et al (2006) quotes a value of 46% on a proximate analysis basis. The oxygen forms part of most of the compounds present in bio-oil. Bio-oil is polar due to its high oxygen content and hence immiscible with non-polar conventional fossil oils (Venderbosch and Prins, 2010; Oasmaa and Peacock, 2010). Oxygen accounts for the low heating value of bio-oil as compared to that of petroleum based fuels, which have insignificant amounts of oxygen. Research has shown that bio-oil have low heating value in the range 14-18 MJ/kg compared to 41-43 MJ/kg for fossil based fuels (Lu et al., 2009). High oxygen content is also responsible for the instability problems associated with bio-oils.

Water present in bio-oil is the sum total of the original moisture in the starting biomass and the water produced as a product of the pyrolysis process (Bridgwater, 2011). The water produced during pyrolysis is as a result of dehydration reactions that occur during the conversion of biomass under pyrolysis conditions (Elliott, 1994; Lu et al., 2009). Of the components of bio-oil, water constitutes the highest percentage by weight. Its value ranges from 15-35 wt% depending on the initial moisture content of feedstock and process conditions (Bridgewater, 2004; Venderbosch and Prins, 2010). Though hard to separate from bio-oil (Bridgwater, 2011; Venderbosch and Prins, 2010), its effect on the quality of the oil can be both positive and negative. Water lowers the heating value of bio-oil and may even cause the oil to separate into phases (Lu et al., 2009; Garcia-Perez et al., 2010; Venderbosch and Prins, 2010). It causes an increase in ignition delay and reductions in combustion rates and adiabatic flame temperatures during combustion (Elliott, 1994; Lu et al., 2009). However, water reduces the viscosity of bio-oils thereby enhancing the flow characteristics of the oil and aiding atomization (Venderbosch and Prins, 2010; Lu et al., 2009; Garcia-Perez et al., 2010). Moreover, it contributes to reductions in emissions during combustion of bio-oils. Water creates more uniform temperature profiles in combustion chambers and this suppresses NO_x formation (Venderbosch and Prins, 2010; Oasmaa and Peacocke, 2010). Also the formation of soot can be hindered by OH radicals from water and can hasten its oxidation (Calabria, et al., 2007).

Viscosity of liquid fuels is very essential especially when the atomization and combustion properties of fuels are considered (Diebold et al., 1995; Lu et al., 2009; Oasmaa and Peacocke, 2010; Venderbosch and Prins, 2010; Garcia-Perez et al., 2010). Generally the viscosity of bio-oils is in the range of 25- 1000 cP at 40^oC and depends mostly on the water content of the bio-oil, which in turn depends on the feedstock and pyrolysis conditions (Bridgewater, 2004; Bridgwater, 2011, Venderbosch and Prins, 2010). Viscosity of bio-oil decreases with increasing moisture content as shown in a study conducted by Sipila et al. (1998). Viscosity also increases with temperature, however above 80^oC the bio-oil tends to be more viscous as a result of aging reactions (Boucher et al., 2000). Besides water, adding polar solvents like methanol, ethanol and acetone has also been found to decrease the viscosity of bio-oils. Studies by Diebold et al. (2005) showed that adding 10-20% by volume of ethanol to bio-oil significantly reduced the viscosity of the oil during storage.

Due to the inefficiency of cyclones, some fine solid particles and ash are always present in bio-oils. The presence of solids is part of the reasons why bio-oils have high viscosities. Char particles catalyse and speed up the aging reactions of bio-oils (Diebold, 2000). This results in a viscosity increase which creates difficulties in pumping and atomization and may even lead to phase separation (Lu et al., 2009; Garcia-Perez et al., 2010; Venderbosch and Prins, 2010). Also these fine char particles tend to agglomerate forming larger particles during storage and may lead to the blockage of fuel injection systems (Oasmaa et al., 1999; Oasmaa and Peacocke, 2010). Almost all mineral matter (ash) in biomass gets sequestered in char particles during pyrolysis, thus making their separation difficult (Ringer et al., 2006). Char particles contains 3-8 times higher ash content than the starting biomass (Lu et al., 2009). Hence a considerable amount of ash ends up in the bio-oil as part of fine char particles thereby posing a detrimental effect on the oil’s quality. Ash is responsible for the catalytic nature of char (Agblevor et al., 1994, Agblevor et al., 1995) and at high temperatures may corrode thermal devices during bio-oil combustion because of the metals found in ash (Morris, 2001). Sodium and potassium metals form low melting compounds that bond to hot metal components leading to their corrosion (Morris, 2001). A considerable amount of the compounds found in bio-oil are organic acids of which formic acid and acetic acid forms the majority. Besides water, the highest concentration of any single substance in bio-oil is acetaldehyde (about 10 wt%) followed by formic acid at about 5wt% and acetic acid at about 3 wt% (Ringer, et al., 2006). Overall the acid content of bio-oil is about 7-12 wt%, resulting in a pH of 2-4 (Lu et al., 2009). This low pH makes bio-oil very corrosive and the corrosiveness is even more aggravated at higher moisture content and high temperatures (Aubin, et al., 1990). Bio-oil corrodes materials like aluminium, mild steel and copper, however, stainless steel and cobalt based materials are resistant (Darmstadt et al., 2004). Some polymers like polyethylene, polypropylene and polyester resins are also not corroded by bio-oils. It is therefore imperative that a careful selection of material is made for use with bio-oils.

Stability, flash point and pour point are very important properties to consider if bio-oil is to be used for fuel applications. Bio-oil is less stable than petroleum based fuels. The instability is caused by viscosity increase during storage resulting from aging reactions. There is also a rapid viscosity increase during heating of bio-oil (Oasmaa and Kuoppala, 2003; Oasmaa and Peacocke, 2010; Garcia-Perez et al., 2010). Thermodynamically, bio-oil production from fast pyrolysis is a non-equilibrium process (Bridgwater, 2004; Ringer et al., 2006; Lu et al., 2009) due to the elevated temperature at which vapours are formed and the short vapour residence times at which they are removed and rapidly quenched. As a result, most of the oxygenated compounds in bio-oils continue to undergo chemical reactions even at ambient temperatures in order to attain equilibrium. These reactions are mainly esterification and polymerization that occur during bio-oil storage (Diebold, 2000). These aging reactions alter the chemical properties of bio-oil, leading to increases in moisture content and viscosity, and possible phase separation (Lu et al., 2009; Garcia-Perez et al., 2010; Venderbosch and Prins, 2010). Molecules of alcohol and organic acids that have high polarity are converted into derivatives of water with extreme polarity and relatively low polar esters by esterification reactions. Polymerization reactions also form large molecules that are less soluble with other compounds in bio-oils, hence increasing the viscosity and leading to eventual phase separation. Fine char particles in bio-oils fuel most of these reactions. They act as catalysts and accelerate these aging reactions. Agblevor et al. (1995) investigated the effect of fine char particle on the viscosity of bio-oil during storage. They observed a remarkable increase in viscosity within a few days and complete solidification within a month when 5 wt% of char was added to a freshly produce bio-oil sample produced from the same biomass.

Aging reactions and instability of bio-oils are also accelerated at elevated temperatures. Initially the oil thickens and phases separate, pyrolytic lignin is then converted into a gummy phase at around 140^{o}C and eventually coke/char is formed from this gummy phase at high temperatures (Oasmaa et al., 1997). Similar observations were made by Boucher et al. (2000) who treated samples of bio-oil at 50^{o}C and 80^{o}C, respectively, and discovered that the bio-oil properties were changed significantly at 80^{o}C, while no significant changes took place in the sample kept at 50^{o}C for a week. Some properties and characteristics of some bio-oil samples compared to that of conventional fuel oil are shown in Table 6.

Table 6: Properties of some Bio-oil Samples compared to Conventional Fuel Oils

Property	**Birch[1]**	**Pine[1]**	**Bagasse[2]**	**Diesel[3]**	**#2 Heavy oil[3]**
Density, kg/m^3	1.25	1.24	1.21	0.85	0.96
Moisture , wt%	18.9	17.0	13.8	0.1	0.1
pH	2.5	2.4	2.7	-	-
Kinematic Viscosity, cSt @ 50°C	28	28	16.4	2.5	351
LHV, MJ/kg	16.5	17.2	17.1	42.9	40.7
Ash, wt%	0.004	0.03	0.05	<0.01	0.03
C , wt%	44.0	48.1	54.6	86.3	86.1
H, wt%	6.9	7.0	6.45	12.8	11.8
N, wt%	<0.1	<0.1	0.73	-	-
S, wt%	0.00	0.02	-	0.9	2.1
O, wt%	49.0	47.0	38.07	-	-
Flash point, °C	62	95	-	70	100
Pour point, °C	-24	-19	-	-20	21

Sources: [1]Ringer et al. (2006); [2]Garcia-Perez et al. (2002); [3]Bridgewater et al. (2002)

From Table 6 it can be seen that the bio-oil samples have densities of about 1.2 kg/m^3, thus making them more dense than the conventional fuel oils which have densities of less than 1 kg/m^3. However, a comparison of heating values show that the conventional fuel oils contain about twice the energy contained in the biomass derived oils. This must be due to the lower moisture content of the conventional fuel oils as opposed to the bio-oils. As discussed earlier, higher moisture content of liquid fuels leads to a reduction in heating value. Pine, birch and bagasse bio-oils contain 18.9wt%, 17.0wt% and 13.8wt% moisture while the conventional fuel oils have almost no moisture.

By comparing the wood (Pine and Birch) derived bio-oils to the bagasse derived bio-oil, one can see that the bio-oil from bagasse is less acidic than the wood derived bio-oils. This therefore makes the bagasse derived bio-oil less corrosive than the wood derived bio-oils. Also it can be seen from Table 6 that bio-oil from bagasse is less viscous than that from the wood feedstock thus making bagasse derived bio-oil more suitable for combustion applications as discussed earlier. Again Table 6 shows that bagasse derived bio-oil have similar heating values as the wood derived bio-oils even though the bagasse derived bio-oil is less dense when compared to the wood derived bio-oils. This is due to the low moisture content and the high carbon content of the bagasse derived bio-oil which makes it to release more energy during combustion.

2.5.1.2.3.3 Uses of Pyrolysis Products

A summary of the applications of fast pyrolysis products is shown in Figure 7 (Bridgwater, 2004; 2011). This includes options for power and heat generation, speciality chemical extraction and a possible upgrade to transport fuels.

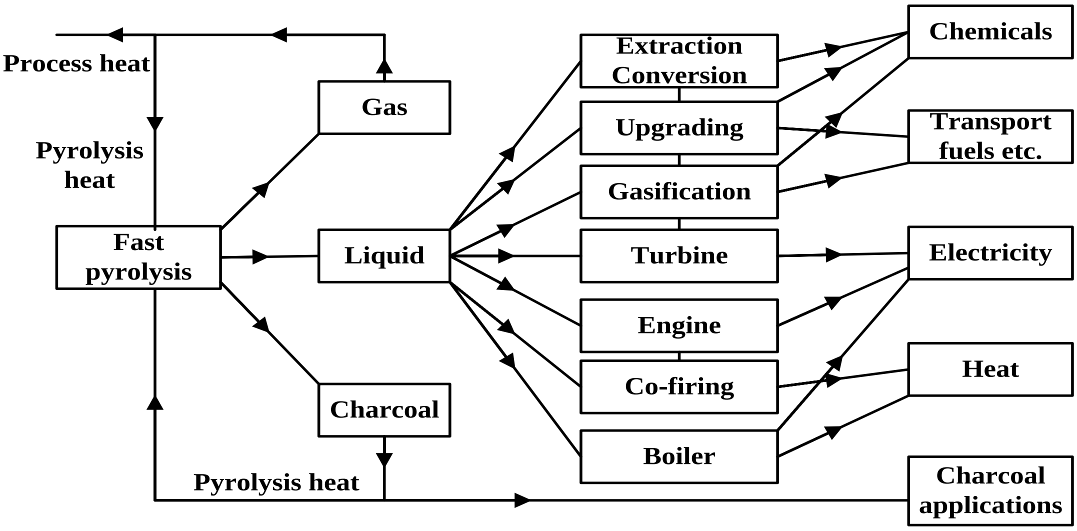

Figure 7: Applications of fast pyrolysis products (Source: Bridgwater, 2011; 2004)

a. Bio-oil (Liquid) Application

i. Combustion

The advantage of liquid products over solids and gases is the ability to easily handle and transport liquids (Bridgwater, 2004). This is very useful when combustion application and equipment retrofitting are considered. The use of solid products such as biomass directly as fuel in oil fired burners will not be possible without major changes to the combustion equipment. However, bio-oil can be effectively used in this equipment with little or no modifications. Furnaces and boilers can operate with several fuel grades such as residual fuels and petroleum distillate, which have several similar properties to bio-oils (Ringer et al., 2006; Venderbosch and Prins, 2010). Thus bio-oil appears to be suitable for combustion in boilers and furnaces and has been tested for combustion applications by several organisations, despite its low heating values. CANMET and MIT conducted combustion tests on bio-oil and concluded that bio-oil does not show significant difference in combustion behaviour compared to #2 fuel oil. Fundamental single droplet test on bio-oil conducted by Wornat et al. (1994) at Sandia National Laboratory also came out with the same conclusion. However one major limitation observed was the high level of emissions (NO_x, CO and particulate matter). This is due mainly to the residual char fines present in bio-oil. Upgrading the quality of bio-oil therefore is a major solution to this problem.

ii. Production of Electricity

The possibility to de-couple fuel production from power generation is an advantage of producing liquid fuels (Bridgewater et al., 1999; Venderbosch and Prins, 2010). This implies that peak power production is possible with a small capacity pyrolysis plant. Diesel engines have higher power generating efficiencies than furnaces and are therefore widely used for electricity generation. Medium and slow speed diesel engines have the ability to run on low quality fuels (Ringer et al., 2006) such as bio-oil. Several diesel engine tests have been conducted using bio-oil and it has shown emission and engine performance results comparable to diesel fuel. A 250 kWe specially-modified dual fuel engine has successfully been operated for about 400 hours by Ormrod Diesels (UK) on bio-oil (Bridgewater et al., 2004). Dynamotive Company in Canada also operates a 2.5 MWe industrial gas turbine which runs on bio-oil (www.dyanamotive .com). There are also reports of successful co-firing of bio-oil with coal in commercial power generation plants in the Netherlands (Wagenaar et al., 2002) as well as at the Manitowac power station in the USA (Free et al., 1996; Venderbosch and Prins, 2010).
One major concern of using bio-oil is its inability to auto-ignite in engines (Solantausta et al., 1993; Bridgwater et al., 2002), which is caused by its low heating value and high moisture content. Other reported problems include corrosion and excessive wear to fuel injection loops of engines (Solantausta et al., 1993; Suppes, 1996). These problems according to Ringer et al. (2006) are due to the acidic nature of bio-oil and the presence of particulate matter (char fines).

It has been found that the auto-ignition delay problem of bio-oil can be solved by the following: (1) addition of nitrated additives to the oil (Solantausta et al., 1993), blending bio-oil with methanol and cetane enhancers and increasing the compression ratio of engines (Suppes, 1996), improving bio-oil quality, and preheating of combustion air to about 55^{o}C (Shihadeh, 1998) and the use of 50:50 diesel to bio-oil blend (Baglioni et al., 2001). The problem of corrosion and wear can be overcome by making careful material selection and improved particulate removal from the bio-oil.

iii. Transport Fuels

The possibility of upgrading bio-oil into transport fuel has the advantage of replacing fossil based transport fuels with an environmentally friendly and renewable source of fuel. Upgrading bio-oil to transport fuel requires the removal of oxygen from the oil and reforming of most of the compounds present in bio-oil (IEA bioenergy annual report, 2006; Ringer et al., 2006; Venderbosch and Prins, 2010). Although this may result in yield loss, because bio-oil contains a high quantity of oxygen (Goyal et al., 2006), it will be balanced by high heating values (Ringer et al., 2006). Upgrading bio-oil is expensive (Bridgwater et al., 1999). Two main methods are been considered to upgrade bio-oil. These are gasification of bio-oil and subsequent synthesis of the syngas and hydro-processing where bio-oil is treated with a hydro-treating catalyst such as Co/Mo or Ni/Mo to remove oxygen/water from the bio-oil and convert it into naphtha-type products (IEA bioenergy annual report, 2006). Another way of upgrading bio-oil is to reject oxygen as carbon dioxide over a zeolite catalyst (Horne and Williams, 1996; Adjaye and Bakhshi, 1995). For a thorough review of the various upgrading technique stated above, one is referred to the review papers of Bridgwater (2011) and Vendermbosch and Prins (2010).

iv. Chemicals

There are over 300 chemicals that have been identified in bio-oil (Soltes et al., 1981). However most of these are found in low concentrations. The potentially higher value of speciality chemicals makes the recovery of these chemicals even at low concentrations practicable (Bridgwater et al., 1999; Bridgwater, 2004).

Chemicals that can be obtained from bio-oil include levoglucosan, hydroxyacetaldehyde, fertilizers, agri-chemicals, polyphenols for resins with formaldehyde, furfural and a variety of food flavourants (Venderbosch and Prins, 2010; Goyal et al., 2006; Ringer et al., 2006). However, the establishment of a market for several of these less common chemicals is required. Speciality chemicals for food flavourings seem to be the option with a viable market opportunity (Bridgwater et al., 1999). There is the need to also devise cheap and efficient separation and refining methods in order to make this more profitable.

b. By-products Applications

From Figure 7 above, the main by-products of fast pyrolysis are gas and char. Due to the endothermic nature of the pyrolysis process (Venderbosch and Prins, 2010), all the gas is combusted to produce the thermal energy required by the process. Some of it is combusted to supply the heat required for the pyrolysis reactor, while the remaining amount is use for biomass feed drying (IEA bioenergy annual report, 2006; Bridgwater, 2011; Brigdwater, 2004; Venderbosch and Prins, 2010).

Depending on how much gas is produced and the heat required, some of the char product may also be combusted together with the gas to generate sufficient thermal energy for the pyrolysis process (Venderbosch and Prins, 2010; Ringer et al., 2006). The remaining char can then be stored and used for other applications. Char can be upgraded to produce activated carbon (Devnarain et al., 2002) and it can also be used directly as bio-char to replenish the nutrient content of soil (Brown et al., 2011).

Activated carbon has found application in many industries because of its excellent adsorbing properties. In industry, it is used to purify, decolourise, deodourise, dechlorinate, detoxicate, filter, recover salts and as catalysts and catalysts supports (Bansal et al., 1988). Its ability to decolourise substances is a beneficial application for the sugar industry. The activated carbon as produced can be used in the sugar refining process to remove colour from raw sugar. For example Lavarack (1997) and Bernado et al. (1997) decolourised raw sugar using powdered activated carbon produced from sugarcane bagasse. Devnarain et al. (2002) also produced activated carbon from sugarcane bagasse and investigated its ability to remove colour from raw sugar. The results showed that the produced activated carbon has a significant potential in the sugar industry.

Carrier et al. (2012) in their studies on bagasse derived bio-char identified some unique properties of bio-char. The addition of bio-char directly to soil, may improve the ability of the soil to retain water, nutrients and other agri-chemicals because bio-char has good absorbing qualities. Bio-char is rich in plant nutrient and minerals and so adds additional nutrients to the soil. Studies by Lehmann et al. (2003) showed that the direct application of charcoal to soil increases the availability of nutrient such as potassium (K) and phosphorous (P) as well as increased nutrient retention for ammonium. Bio-char also acts as liming agent thereby assisting in the control of soil acidity.

2.5.1.2.4 Implementation of Pyrolysis in a Sugar Mill

In recent years, several researchers have focused on finding alternative uses of sugarcane residues as energy sources. Among these is the use of sugarcane residues as fuels in the sugar industry to generate steam and electricity to run the industry. It has been concluded that sugar industries need to undertake development in multi-product biorefineries, instead of focusing only on producing sugar, to remain economically sustainable. This therefore calls for the efficient use of the waste generated in the industry. Currently research has focused on making the sugar industry as efficient as possible in terms of energy and this has resulted in most sugar industries generating enough electricity for sale to the grid as in Mauritius (Mbohwa, 2003).

Among the waste generated in the sugar agro-industry are bagasse and SCAR (sugar cane agricultural residues, e.g. leaves and tops). Font (2000) in his studies on the energy potential of sugarcane in Cuba estimated that 5.69 tons of biomass residues (wet basis) are generated per ton raw sugar produced, with SCAR and bagasse accounting for 42% and 45%, respectively. Table 7 shows the yield of residues generated during the production of raw sugar.

Table 7: Sugarcane Residues Yield per Metric Ton of Raw Sugar Produced

Sugarcane residue	**Yield, ton/ ton of raw sugar**
Bagasse (50% moisture)	2.57 ton
Trash at cleaning centre (50% moisture)	0.57 ton
Trash remaining in field (50% moisture)	1.85 ton
Molasses	0.4 ton
Filter cake	0.3 ton

Source: Font (2000); taken from Pippo et al. (2007)

Currently, bagasse and, in some cases, some amount of SCAR is combusted in the sugar industry to generate steam and electricity to run the sugar milling and refining processes, and also as a means of disposing of these residues. However, the full energy potential of these residues has not been exploited. On energy basis, a ton of bagasse (50% mill- wet basis) is equivalent to 1.6 barrel of fuel oil (Pippo et al., 2007). With research focusing on energy integration and minimal use of energy in the sugarcane processing industry, the implication is that more surplus bagasse can be produced, thereby increasing the amount of energy available in waste generated. Storing large volumes of bagasse and SCAR is costly and poses environmental threat due to their low bulk densities (large volume required for storage) and self-combustion properties (Lavarack, et al., 2002; Pippo et al., 2007). Hence storing these residues for use during off-season for steam and electricity production is not a suitable option and a more practicable solution to this is needed.

Pyrolysis has the ability to convert biomass such as bagasse and SCAR into high density fuels that can be stored for use later off-season to run the sugar processing plant, or to provide additional income from value-added energy-products. The pyrolysis process by which these products are manufactured also generates large quantities of thermal energy, which may be used to produce steam and electricity to run the sugar mill. Not much has been done with regards to the implementation of pyrolysis in the sugar industry. Pippo et al. (2007) investigated the possibility and economics of implementing a 3 ton/h fast pyrolysis model in a sugar factory in Cuba. This was done to find an alternative source of energy for use by the sugar industry during the off-season. Conclusions drawn showed that the efficient use of sugarcane residue and the on-site implementation of fast pyrolysis in a sugar mill alongside energy integration measures in the sugar making process is cost effective and could result in an anticipated excess of 80 KWh_e/ton of cane per season or 6 million tons of bio-oil for use off-season in a milling season of 4 million tons of raw sugar.

The integration of fast pyrolysis models in the sugar mill process offers several advantages. Some of these advantages as listed by Pippo et al. (2007) are as follows:

1. The ability of the existing mill infrastructure to absorb the technology of fast pyrolysis. This implies that not much resource is needed for infrastructural development.
2. Pyrolysis provides a cost effective and safe alternative to storing sugarcane residues. It converts low energy dense residue to high energy dense bio-oil which is considered innocuous in terms of CO_2 emissions, hence environmentally friendly.
3. Bio-oil can be handled by existing infrastructure used for transporting and distributing conventional fuels.
4. Transporting bio-oil is easier than biomass and so can be transported to remote locations for use as domestic fuel.
5. About 11 times more energy is stored in bio-oil in the same unit volume of biomass.
6. The ability to store bio-oil implies that its production can be separated from the power generating cycle. This ensures that it is only used when required and in the right quantity.

Other benefits include the following:

a) The possibility of extracting valuable chemicals from the oil for sale, adding to the revenue of the mill.

b) The char produced as a by-product of pyrolysis may be suitable for use directly on sugarcane plantations to enrich the fertility of the soil. Such biochar has a high absorbent property and so when added to soil, may improve the soil’s ability to hold water and nutrients. Besides it is rich in ash and minerals which are released into the soil to improve on the soil’s fertility.

c) The possibility of upgrading char into activated carbon. This can be used in the sugar refinery as a colour adsorbent to remove colour from raw sugar during the refining process.

Despite these benefits, the unstable nature of bio-oil is a major challenge. Another factor is the endothermic nature of the conversion process (pyrolysis). However, though the process is endothermic, requiring the input of energy, this energy can be recovered during pyrolysis product recovery through the implementation of effective heat recovery systems, which can then be used to raise steam for use in the sugar mill and also generate electricity.

Some important particularities for using fast pyrolysis module at sugar mills include, but are not limited to, the following (Pippo et al., 2009; Pippo et al., 2007):

i. Increases in profit margins since feedstock may not need to be transported.

ii. The possibility of adjusting the pyrolysis module to the milling capacity of the sugar factory. Hence there is flexibility.

iii. The possibility of a cluster of sugar mills making use of a centralised fast pyrolysis facility.

iv. There is the possibility of carrying out mobile pyrolysis on small scale in surrounding plantations and bringing the product to a centralised storage unit.

v. Also the normal operation of the sugar mill would not be affected since the bio-oil production can be done on request.

2.5.2.4.1 Bagasse and SCAR as raw materials for pyrolysis in the sugar mill: availability and properties

Bagasse is the fibrous remains of sugarcane stalks after crushing and extraction of juice. Like any other biomass, it is composed of cellulose, hemicelluloses and lignin. It is estimated that approximately 240 kg of bagasse (50% moisture) is produced per ton of sugarcane processed (Dias et al., 2009) or 1.25 kg of dry bagasse is produced per kilogram of sugar produced (Botha and Von Blottnitz, 2006; Singh et al., 2007). The South African sugar industry produced an average of 1.35 kg dry bagasse per kilogram of sugar during the 2011/2012 milling season (S. Davis, SMRI, personal communication). Even though bagasse have well known energy properties, it is currently utilized as fuel in co-generation systems, where it is often burnt with low efficiency boilers, designed to raise enough steam and electricity for the production process, while leaving little or no surplus of bagasse – boilers are therefore used to dispose of bagasse, a by-product of the sugarcane milling process. However, with the current trend of most mills moving towards installation of more efficient cogeneration systems, greater quantities of bagasse are being produced and the implementation of energy integration systems in sugar mills will even increase the quantity available. Hence, there is enough bagasse available for use as feedstock in producing high valued products from pyrolysis alongside electricity generation for the sugar industry.

Bagasse as energy feedstock has varying composition and heating values. Its characteristic property depends mainly on the climate, type of soil on which cane was grown, cane variety, harvesting method, amount of cane washing and the efficiency of the extraction plant (Janghathaikul and Gheewala, 2004). Generally the high heating value (HHV) of bagasse is about 7-9 MJ/kg at moisture content of 45-50%, as produced from most milling processes. Oven dry bagasse has HHV of about 18

MJ/kg (Garcia-Perez et al., 2002). The proximate and the ultimate analysis of sugarcane bagasse and SCAR can be seen in Table 8 and Table 9 respectively.

Table 8: Sugarcane Trash and Bagasse Proximate Analysis and Heating Value (Macedo et al., 2001)

Sample	**Proximate analysis, wt%**				**Higher heating value[1], MJ/kg**
	Ash[1]	**Volatile matter[1]**	**Fixed carbon[1]**	**Moisture**	
Dry leaves	3.3	85.8	10.8	11.3	17.4
Green leaves	3.2	85.2	11.7	66.7	17.4
Tops	4.3	84.1	11.6	82.5	16.3
Bagasse	3.0	83.0	14.0	50.0	18.0

[1] Dry basis

Table 9: Sugarcane Residue Ultimate Analysis (Fernandez et al., 2004: in Pippo et al., 2007)

Sugarcane residue	**Ultimate analysis % dry matter**					
	C	**H**	**O[1]**	**N**	**S**	**Ash**
Bagasse	47.4	7.2	40.69	0	0	4.71
SCAR	46.0	6.6	41.7	0	0	5.65

[1] By difference

From Tables 8 and 9 above, it is seen that SCAR is similar to sugarcane bagasse in terms of energy content and chemical composition. The average heating value of SCAR is about 17 MJ/kg on dry basis. Despite this value of SCAR it is usually burnt in the field to facilitate the cane harvesting process or left on the field to decompose and serve as manure; supplying nutrients to the soil. In terms of availability, the amount of SCAR from sugar cane harvesting depends on the cane variety, age, and soil and weather conditions. However, the actual quantity depends on the harvesting method used ie., manual or mechanical (Mecedo et al., 2001). About 35 wt% SCAR is obtainable per ton of mill-able cane (Pippo et al., 2007). Therefore with South Africa producing about 22 million tons of sugarcane, 7.7 million tons of SCAR is produced annually. The similarity of SCAR to bagasse in terms of energy makes it a valuable resource that can be used as a clean source of energy. One way of making use of SCAR is its conversion into energy carriers for use during off-season.

One important feature to consider as far as using bagasse and SCAR as feedstock for pyrolysis is concerned is their high ash content (see Table 8 and 9). Inorganic minerals (ash) usually sequester in char during biomass pyrolysis, and depending on the efficiency of the solid separation system used in the pyrolysis plant, some of this char may end up in bio-oil. This affects the quality of bio-oil and may have detrimental effects on equipment that may use this oil. Ash and alkali metals in ash aid slagging and fouling formation (Pippo, et al., 2007). Studies by Devnarain et al. (2002) reported an ash content of between 1.8-5.4 wt% for South African sugarcane bagasse (2 wt% moisture content). The high mineral/ash content of bagasse is mainly due to contamination from soil particles rich in inorganic components. Ash content of SCAR also depends largely on the method of cane harvesting. SCAR can be contaminated with soil particles to a much higher degree than bagasse and so the reason for its higher ash content. Washing of sugarcane residues is a possible way of reducing ash/mineral content. More than 80% of K and Na alkali metals can be removed by the simple washing of sugarcane residues (Turn et al., 2002; Keown et al., 2005). This means that the juice extraction step of the sugar milling process in which imbibition water is added to the sugarcane to aid in maximum sucrose recovery, also serves as a way of reducing the ash content of bagasse. This is because most of the soil particles that might be attached to the outer part of the sugarcane stalks is removed from the cane fibres (bagasse) into the raw juice. Because SCAR does not take part in the milling process it may require an additional washing step before its use as feedstock for pyrolysis at the sugar mill. Pre-treatment of SCAR with low concentration HCl solution could also reduce the alkali content (Pippo et al., 2007). Another possible way of using SCAR for pyrolysis while avoiding increase in ash content of the resulting bio-oil is to mix it with bagasse at reasonable proportions.

CHAPTER THREE

Design Basis and Approach

3.1 Introduction

This chapter provides information on the basis and assumptions applied for the conceptual design of the biomass-to-energy conversion processes (BMECP) modelled in this study. The BMECP is the so-called "energy island" that provides the energy requirements of a sugar mill using available biomass. In this thesis possible options for use of pyrolysis, in comparison to the efficiency of combustion, as a central process for the "energy island" were considered. All process models were developed using the Aspen Plus$^{\circledR}$ simulation software. As stated already in chapter one, two BMECP processes (combustion and pyrolysis) were investigated in this study and the path followed for the conceptual design of the various flow sheet configurations is as presented in Figure 8. It includes the choice and development of appropriate process flow diagrams and operating conditions, which was done after a careful study of literature as well as in consultation with industry partners at the SMRI (Steve Davis and Richard Loubser - personal communication). This was then followed by the estimation of the mass and energy balances of the various process configurations using Aspen Plus$^{\circledR}$ which were then used for the sizing of process equipment and subsequent economic evaluation of various scenarios using Aspen Icarus$^{\circledR}$ and information from other databases. The basis and the assumptions made for the sizing of equipment and economic analysis are discussed in details in the economic modelling section (Chapter 5) of this thesis.

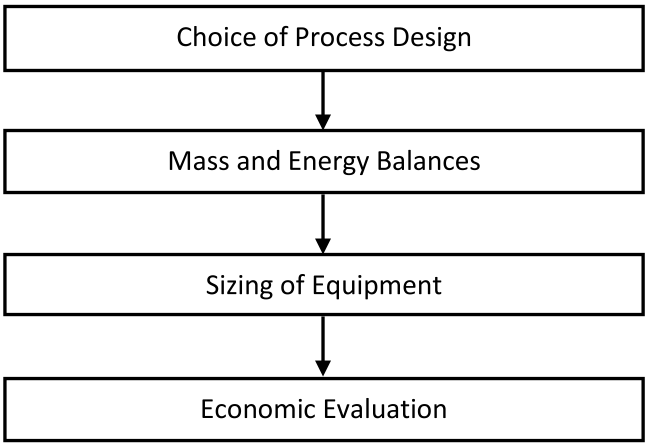

Figure 8: Conceptual Design Flow Path

3.2 General Overview of BMECP and Scenario Selection

Figure 9 below shows the schematic process block flow diagram (PBFD) of a BMECP plant integrated to a sugar mill. The BMECP was considered in this study to be a stand-alone power plant dedicated to the production of electricity, steam and other energy products (bio-oil and biochar – in the case of pyrolysis). The energy demands of the sugar mill in the form of process steam and electricity (it is assumed that all prime movers in the sugar mill run on electric motors) are therefore met by the BMECP plant, while any excess energy products are sold to generate revenue. It was thus assumed that the BMECP plant is located near a sugar mill and that the sugar mill in turn supplies it with biomass (bagasse) feed and condensate water for its operation. Facilities for water treatment and waste treatment were therefore assumed to be already in existence in the sugar mill and thus the process equipment for these utilities were not included in all the process models developed in this study.

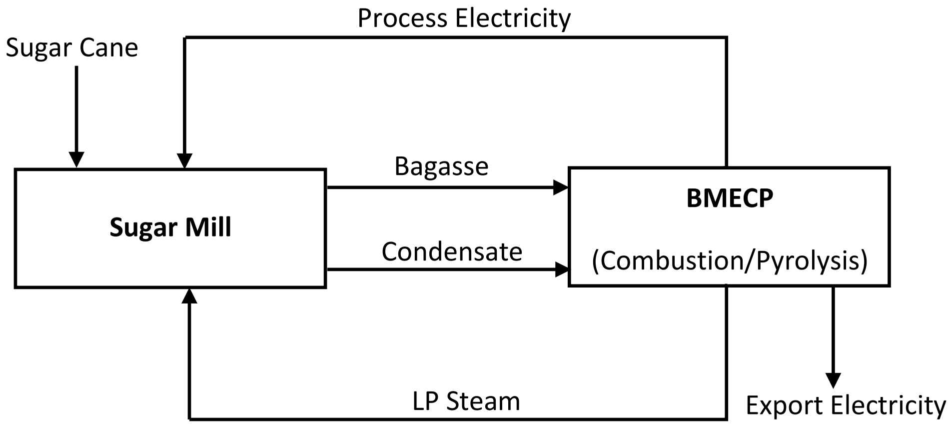

Figure 9: Schematic Block Flow Diagram of a BMECP lant

3.2.1 Choice of Scenarios

Three main scenarios of the BMECP were selected in this study as the technological pathways for the conversion of sugar mill biomass into energy and energy products. The first scenario referred herein as Combustion BMECP (Figure 10), models the current technological pathway used in the sugar industry for the generation of electricity and steam to run the sugar milling process. Here sugar cane bagasse from the sugar mill is fed to a biomass combustor/steam-turbine cycle to cogenerate electricity and LP steam for the sugar mill.

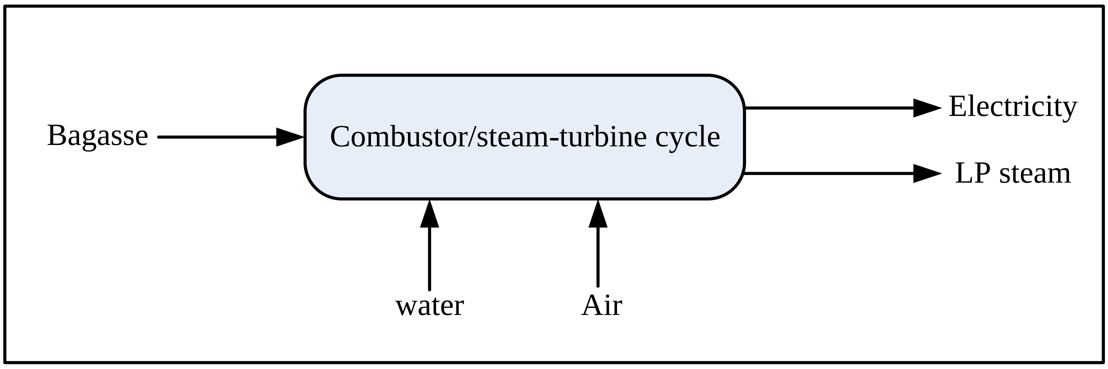

Figure 10: Combustion BMECP BFD

In the second scenario all bagasse from the sugar milling process is fed to a fast pyrolysis plant where it is converted into pyrolysis products (bio-oil and biochar). Part of the pyrolysis products is combusted to supply the energy for pyrolysis and also to cogenerate electricity and steam in a steam-turbine cycle for sugar mill operations. This scenario was referred herein as Pure Fast Pyrolysis BMECP (Figure 11).

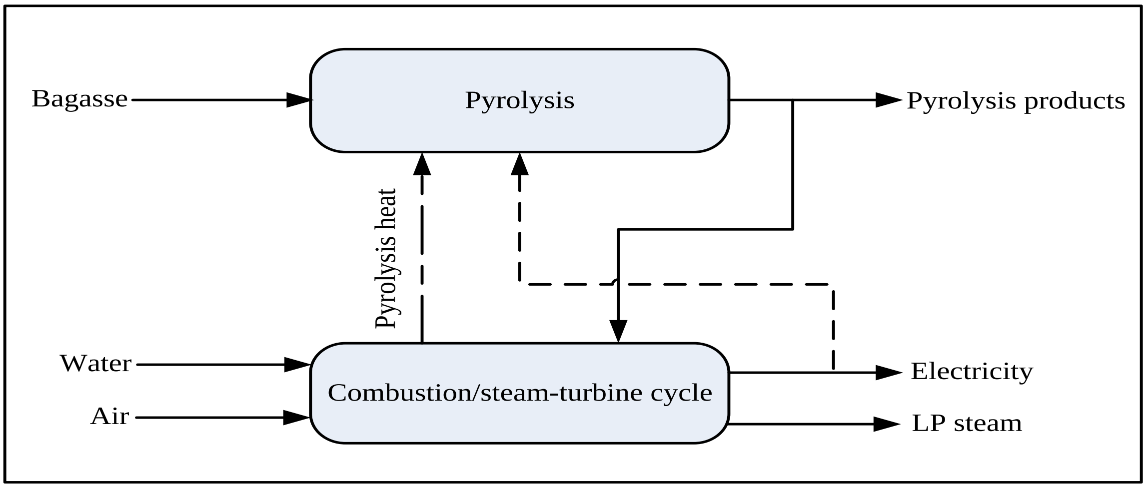

Figure 11: Pure Fast Pyrolysis BMECP BFD

The third scenario also referred herein as Partial Fast Pyrolysis BMECP (Figure 12) is a variant of the Pure Fast pyrolysis process. Here bagasse feed from the sugar mill is split into two fractions. One fraction, which is sufficient to provide the energy requirement of the sugar mill, is combusted in a biomass combustor/steam-turbine cycle to generate steam and electricity for the sugar mill operation. The remaining fraction is fed to a fast pyrolysis plant to produce bio-oil and biochar as final products. No part of the pyrolysis product is combusted. Heat for pyrolysis is also provided by the biomass combustor.

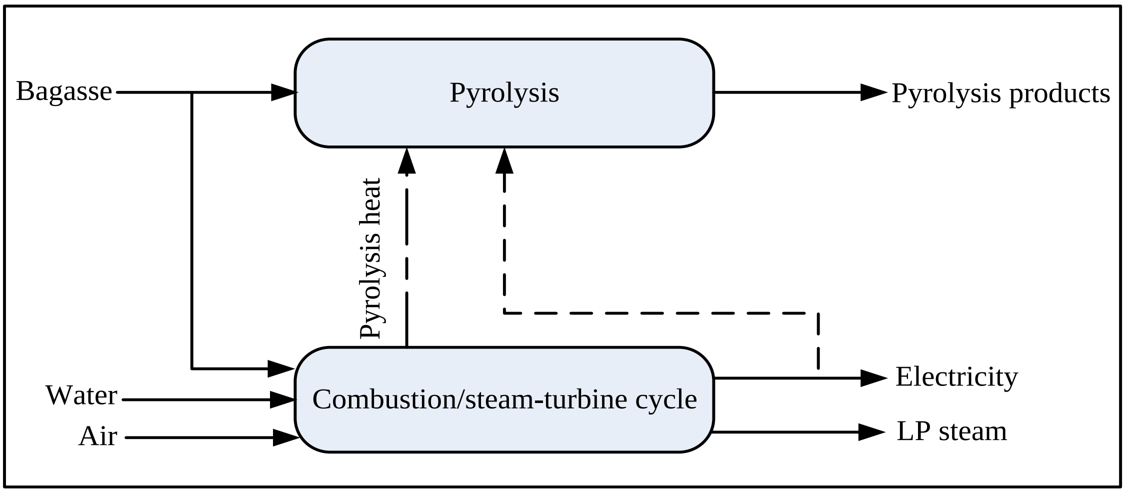

Figure 12: Partial Fast Pyrolysis BMECP BFD

For this study, two cases of a sugar mill were considered for each of the BMECP scenarios. These include (1) a sugar mill operating on 50% mass of steam to the mass of cane crushed (less efficient mill) and, (2) a sugar mill operating on 40% mass of steam to the mass of cane crushed (more efficient mill). The selection was done in close consultation with the SMRI. The 50% steam on cane was chosen to reflect the current state of steam consumption in most South African sugar mills. The 40% steam on cane was chosen as a hypothetical case of an efficient sugar mill in order to determine the effect of process steam efficiency on the energy production capacity of a BMECP plant and the economics thereof. It is envisioned that sugar mill efficiency of about 38% steam on cane is achievable in the near future as sugar mills across the world implement energy integration measures in the mills to cut down on steam usage (Sharma & Peacock, 2008; Pippo et al., 2009). For both sugar mill efficiencies, models of each BMECP scenario were developed to assess the ability of both the Combustion (current technology used in the sugar industry) and the Pyrolysis-based BMECPs in supplying the energy demands of the mill and possible generation of excess energy/energy products.

A 50% steam on cane sugar mill using combustion technology as the BMECP and a 30bar pressure back pressure steam turbine (BPST) turbo-alternator cogeneration system running on 50% moisture bagasse feed as fuel was chosen as the base case scenario in this study. This models the current BMECP conditions used in South African sugar mills. For the Combustion BMECP, process models were developed to study the effect of boiler operating pressure and bagasse moisture content on the production of steam and electricity for each of the sugar mill efficiency cases stated above. Pressures of 40bar, 63bar and 82bar and bagasse moisture contents of 48% and 46% were considered, in addition to the base case of 30bar boiler pressure and 50% moisture content, respectively. The selection of the above pressures is in accordance with the operating pressures applied in most biomass operated power plants around the world. Though higher pressures could have been chosen, a study by Mbohwa (2003) shows that at steam pressures beyond 160bar, there is a drop in process efficiency due to the severe decrease in latent heat of steam, which subsequently decreases the rate of heat transfer and hence the electricity production. Also a large percentage of the boiler feed water used in the sugar mill is supplied from returned mill condensate, which might have high levels of total dissolved solids (TDS) and silica. A study by Avant-Garde Engineers and Consultants in India (www.avantgarde-india.com) have shown that these TDS and silica gets carried over as vapour together with the steam at higher boiler operating pressures thereby affecting the purity of the steam, which in turn affect the normal operation of downstream process equipment such as the steam turbine due to depositions on the blades of the turbine leading to corrosion problems. Magasiner et al. (2001) in their study also found that there is a buildup of carbon and ash on the water side of boiler tubes when operating at higher pressures. This might lead to problems such as reduced heat transfer rate, blockage, slagging, and eventually tube failure due to corrosion. Another reason is that operating at higher pressures comes with an extra cost of the power generating system as materials with special properties are needed, hence, the selection of the above pressures. Also although bagasse drying was not considered in this work for the combustion BMECP technology, it was assumed that the sugar mill can achieve the 46% and 48% bagasse moisture contents by improving the efficiency of bagasse dewatering and drying mills (Hugot, 1986; Steve Davis, personal communication). From literature, it is shown that at below 46% biomass moisture content, other factors besides moisture play a critical role during biomass combustion, affecting the performance of the system. Drier fuels leads to increased high flame temperatures causing problems such as fouling of heat transfer surfaces, grate level slugging and sometimes even explosion (Mugadhi, 1999), thus the choice of 46% as the lowest moisture content for this study. The two pyrolysis-based processes were modelled for only the 50% bagasse moisture content.

3.2.2 General Overview of Combustion BMECP

The model as built in Aspen Plus® was based on the Rankine cycle of a Combined Heat and Power (CHP) plant, which was modified to accommodate steam extraction at sugar mill conditions of 2bar pressure and 120^{o}C. It consists mainly of the biomass combustor; steam generator (boiler) and turbo-alternator (Figure 13).

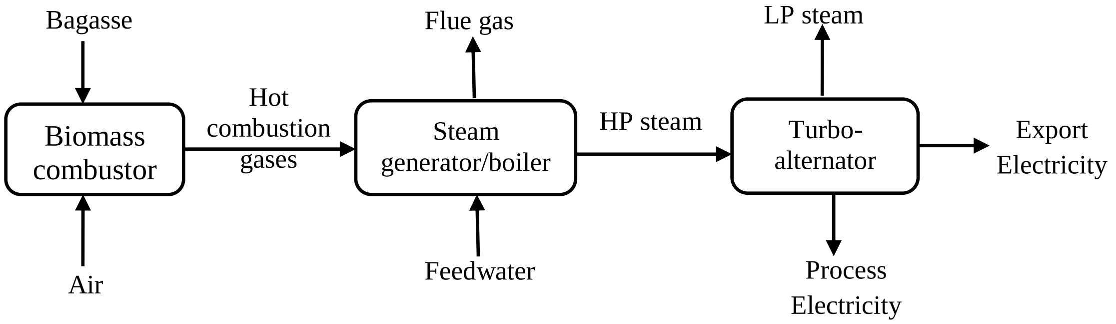

Figure 13: Schematic representation of the combustion process

In this process, bagasse fuel (at given moisture content) and combustion air are fed to the biomass combustor. Hot combustion gasses generated after combustion are used to generate steam in the boiler/steam generator through indirect heat exchange with boiler feed water. Prior to exiting into the atmosphere, the flue gases are cleaned to remove ash and particulate matter. It was assumed in this study that all the energy required for steam production is solely supplied by the bagasse feed from the sugar mill; hence no external heat source to the biomass combustor was included. Combustion air is preheated with flue gas in the air pre-heater to raise its temperature to 250 ^{o}C prior to being fed to the combustor. Air was supplied in excess (40-50% excess) to ensure complete combustion and also to make sure that the percentage volume of oxygen in the flue gas is at least 6% so as to meet environmental standards as set out in the Department of Environmental Affairs and Tourism Air Quality Act (2008).

In the steam generator, hot combustion gasses from the combustor exchanges heat with boiler feed water through indirect-contact heat transfer, resulting in the generation of high pressure (HP) steam. The quantity of HP steam produced depends on the heat available as well as the quality (superheated conditions) of the HP steam required. To ensure efficient generation of steam, the feed water to the boiler is pre-heated. All heat for the steam generation is supplied by the hot combustion gasses.

HP steam from the boiler is expanded in a turbo-alternator, Back-Pressure Steam Turbine (BPST) or Condensing Extraction Steam Turbine (CEST), depending on the pressure of the steam, to produce electricity and an amount of low pressure (LP) steam is extracted and sent to the sugar mill for use as process steam.

3.2.3 General Overview of Pyrolysis-based BMECP

A general overview of the two pyrolysis-based processes is shown in Figure 14. It consists of the following subsection: pre-treatment, pyrolysis, condensation and oil recovery and a combustion/steam turbine cycle. A detailed description of each flowsheet section as modelled in Aspen Plus® is given in chapter four.

The pre-treatment section involves the drying of bagasse feedstock from a given moisture content as received (50%, 48% or 46% - depending on the mill condition), to between 7-10% moisture content using preheated process air; this is then followed by grinding to reduce the particle size to about 2mm, which is the recommended particle size for fast pyrolysis. Drying of the feedstock is done in order to reduce the final moisture content of the bio-oil produced. Higher moisture content reduces the heating value and quality of bio-oil (Bridgwater and Peacocke, 1999; Ringer et al., 2006).

Dried bagasse is sent to the pyrolysis reactor where it undergoes pyrolysis at 500°C temperature and atmospheric pressure yielding char, condensable biocrude components and non-condensable gases. Char is separated from the condensable and non-condensable gases using a cyclone that follows directly after the reactor and a part of the char is combusted together with the off-gas from the reactor to supply the energy needed to run the pyrolysis process. Some reactor gas is recycled for use as a fluidising medium in the reactor.

In the condensation and oil recovery section, gas product exiting the cyclone is immediately cooled to recover bio-oil. Cooling is first done using water followed by the use of air in condensers. This results in the generation of some steam as well as hot air, which is then used for bagasse drying during pre-treatment. An oil scrubber following the condensers recovers almost all the remaining bio-crude components in the non-condensable gasses. Vapour phase exiting the scrubber is then sent to an electrostatic precipitator for further oil recovery. All bio-crude recovered is sent to the product recovery section where it is cooled to room temperature and stored. Here, a small fraction is recycled to the scrubber to enhance oil recovery.

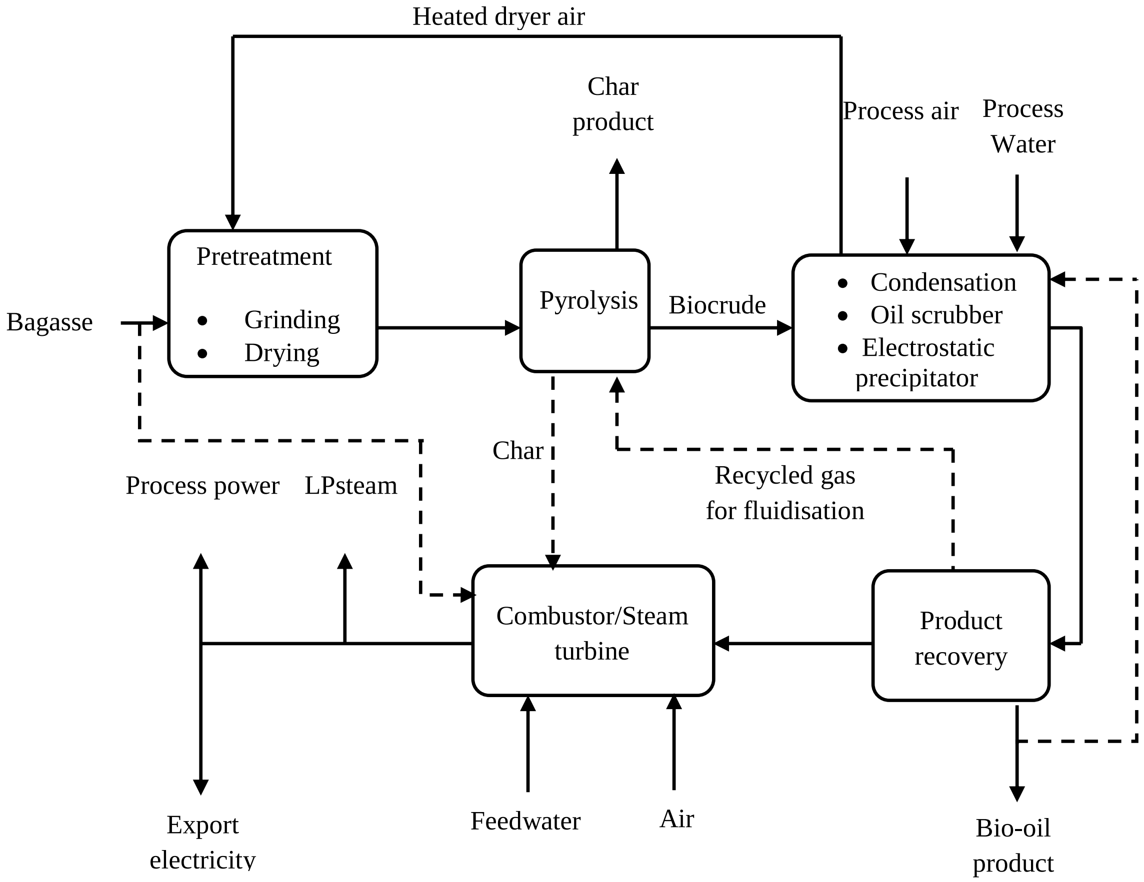

Figure 14: Schematic representation of the pyrolysis process

After the recovery of bio-crude components, the remaining non-condensable gas is sent to the combustion and steam turbine section where a part is removed and sent to the pyrolysis reactor as a fluidizing medium and the rest combusted together with a portion of biochar and/or bagasse to generate HP steam. Air is supplied in excess (40-50% excess) to ensure complete combustion. The HP steam is then expanded in a steam turbine to generate electricity with LP steam extraction for use by the sugar mill as process steam. Steam produced during quenching is added to the one produced in this section before electricity generation. How much bagasse is combusted together with the non-condensable gas depends on the type of pyrolysis BMECP configuration considered.

For a pure pyrolysis BMECP system, no bagasse is burnt while bagasse is combusted together with non-condensable gas in the scenario which considers a combination of both combustion and pyrolysis to supply the energy needs of the sugar mill in a single BMECP plant.

3.3 Development of Mass and Energy Balances

Depending on the BMECP type and the choice of scenario considered, process flow sheets (as indicated in appendix A) were built in Aspen Plus® and the necessary operating conditions of unit operation blocks/equipment and input streams as well as product specifications entered into the program to generate the overall mass and energy balances of the process. All process flow sheets were modelled to accommodate the highest level of energy integration possible in a particular process in order to minimise the waste of energy and guarantee improved process efficiency. Some of the assumptions adopted in developing the various Aspen Plus® process models are discussed below.

3.3.1 Bagasse Throughput and Composition

All process models developed in this study were based on a 300 ton of cane/h throughput. This value was calculated from data supplied by the SMRI. It reflects the average cane crushing of South African sugar mills (14 mills in total) for the 2009/2010 milling season, estimated to be 293.33 tons of cane/h (Smith et al., 2010). However, the figure 293.55 was rounded up to 300 tons of cane/h in order to simplify the models. This was also done so as to satisfy the milling capacity value (7000-10000 tons of cane/day) of a medium sized sugar mill (Pippo et al., 2009). Based on the assumption of 270kg bagasse/ton of cane crushed (Garcia-Perez et al., 2002), the 300 tons of cane/h yields 81tons of bagasse/h (50% moisture as received from the sugar mill) and this was set as the feed input capacity for all the various BMECP process scenarios modelled.

Table 10: Bagasse Composition

Component	**Value, %**	
	This work (Hugo, 2010)	**Garcia-Perez et al. (2002)**
Moisture	50	50
Lignin, (dry basis)	25.5	20-27
Cellulose, (dry basis)	40.6	35-50
Extractives, (dry basis)	7.5	8-12 (for total extractive and ash)
Ash, (dry basis)	3.6	
Hemicelluloses, (dry basis)	22.8	20-30
Elemental Composition		
C	50.3	
H	6.3	
O	43.1	
N	0.3	
S	0.07	

To effectively simulate the process for the conversion of bagasse to energy and energy products, it is important to know its composition. Good data on bagasse compositional analysis is thus very vital. Table 10 above shows the values adopted for this work. These are average values computed from experimental result obtained from bagasse compositional analysis work done at Stellenbosch University (Hugo, 2010) and compares very well with values reported in literature (Garcia-Perez et al., 2002).

3.3.2 Physical Property Data and Stream Component Specification

Choosing an appropriate physical property data for process stream components is essential when doing process modelling/simulation. This is done to ensure accurate prediction of the thermodynamic properties of components which in turn will produce a reliable simulation result.

For the modelling work done in this study, the physical property data of most of the stream components are already found in the databank of the Aspen Plus® simulation software used. However, due to the complex nature of bagasse, the physical property data (such as density, heat capacity, heat of formation) for most of its components could not be found in the Aspen Plus® property databases. Properties for these components were therefore taken from the in-house Aspen Plus® property database developed by the National Renewable Energy Laboratory (NREL) (Wooley & Putsche, 1996). Lignin, cellulose, extractives, and hemicellulose were the components taken from the NREL database. The hemicellulose component was further split into xylan, mannan, galactan and arabinan to simplify the model. Further information on these biomass components as used in the process models is presented in Appendix B

3.3.3 Stream Component Specification

Components used in the simulations are classified as either MIXED components or conventional inert solid (CISOLID) components in Aspen Plus® using stream classes. CISOLID is applied to solids components with a defined molecular mass that do not participate in phase and/or equilibrium calculations. All MIXED components take part in phase and/or chemical equilibrium calculations (Aspen Technology). In this study, the solid part of the bagasse feed was modelled as CISOLID, while all the other components (moisture) were modelled as MIXED.

3.3.4 Utilities

As indicated in Figure 9 (section 3.1), the BMECP plants modelled in this study were assumed to be stand-alone power-and-steam plants located close to a sugar mill, using sugar cane bagasse and process condensate from the mill, while providing the mill with electricity and LP steam. All electrical energy used by the BMECP equipment, such as pumps and compressors, are met by the plant itself, hence water for use as boiler feed water is the only utility that was sourced from outside the BMECP plant.

It was assumed that condensate from the raw sugar house is used as boiler feed water. The temperature of LP steam as used at the raw sugar house is usually within the range of 120^{o}C and 130^{o}C. Therefore from a thermodynamic point of view, the temperature of the condensate from the mill should be around the same temperature range. However, due to activities such as vapour/steam bleeding from multiple effects evaporators for heating purposes in heaters and vacuum pans, condensate temperature is usually below the 120 – 130^{o}C range (mostly below 100^{o}C - S. Davis, SMRI, personal communication). Condensate temperature is further reduced considering the fact that not all condensate would be recovered as "pure condensate" for use as boiler feed water, hence the need for the addition of make-up water. Adding make-up water to the condensate causes its temperature to further fall below the 120^{o}C – 130^{o}C temperature range. The exact temperature reached depends on the quantity of make-up water added. For this reason, the boiler feed water temperature was set at 80^{o}C in this study.

3.4 CO_2 Savings/Environmental Impact

To determine the impact of each BMECP technology (combustion and pyrolysis) on global warming, savings in CO_2 (main greenhouse gas) were estimated for each process model developed in Aspen Plus®. For this work, a 'first hand' method rather than a full lifecycle assessment (LCA)/environmental impact assessment (EIA) was used to estimate the savings in CO_2. A full LCA is a complete study on its own as done in the work of Botha and Blottnitz (2006) and thus is beyond the scope of this work, hence the choice of this 'first hand' method.

This method assumes that all electrical power/LP steam/energy products generated by a particular BMECP could have been produced by a similar power generating plant running on fossil based fuel as its source of energy input (Larson et al., 2001). The amount of fossil based fuel required to generate an equivalent amount of electricity was thus calculated taking into account the conversion efficiency of such a plant after which the quantity of CO_2 produced by such a facility was estimated as the potential savings in CO_2 for a particular BMECP scenario.

In this study a coal fired power plant was considered as the alternative to power generation from biomass in the BMECP plant. This assumption is validated by the fact that about 93% of current electricity generation in South Africa is from coal (Eberhard, 2011). Hence CO_2 savings was calculated based on the assumption that the bagasse electricity displaces electricity produced from coal. Also, the amount of coal equivalent to all thermal products (LP steam and/or bio-oil and biochar) was calculated for each scenario using the respective heating values of each thermal product and coal, after which the CO_2 savings was calculated and added to that obtained from electricity generation. The coal assumed in this study is a South African bituminous coal with an average heating value of 26 MJ/kg (Eberhard, 2011).

The average efficiency of a coal fired power plant is estimated to be around 30% (European Association of National Metrology Institute [EURAMET], 2009). Working at this efficiency, it was estimated from literature that approximately 0.5 kg of coal is required for each kilowatt hour of electricity produced (Mbohwa and Fukuda, 2003; EURAMET, 2009), which results in CO_2 emissions of 2.62kg (Mbohwa and Fukuda, 2003) and 2.325kg (EURAMET, 2009) per kilogram of coal. For consistency, the average value of 2.47 kg CO_2/kg coal was used for this study and all CO_2 savings were thus estimated based on this value. For calculations, the CO_2 emitted by the BMECP plant itself was not taken into account. This is based on the assumption that sugarcane bagasse is a renewable fuel and so all CO_2 produced by the BMECP plant during power/energy products generation would be reused by sugarcane plants for photosynthesis during the sugarcane planting season (Basu, 2010). Although some fossil fuel is consumed during the growing, harvesting and transportation of sugar cane, the carbon released from this source was also not taken into account as it would have been in a full LCA.

3.5 Process Energy Performance

The energy efficiency of a process is a measure of how well a process converts the energy content of the feed input to useful energy in the final products produced from the process. It is the ratio of the energy content of products to the feed energy supplied to the process. Estimating process efficiency allows one to compare different process technologies/configurations that can be employed for the conversion of raw material/s into finished product/s which results in informed decision making.

Several methods are used in calculating the energy efficiency of a process. For a process that is involved in the generation of both thermal and electrical energy (CHP) for end-user applications, two main methods are applied in estimating the process energy efficiency. These are (1) the electrical efficiency and, (2) the overall process efficiency. Since the objective of the BMECP plants modelled in this study is to produce both electricity and thermal energy from bagasse, both efficiencies were calculated and used to compare the various process model scenarios. The definitions of these efficiencies and their equations thereof are given below (Mani et al., 2010).

The electrical efficiency (see Equation 1) is defined as the ratio of the net electric/power output generated by a process to the total energy contained in the input fuel. Net electric/power output is the difference between the gross electricity generated by the process and the sum of all electrical power used by the process to run process equipment such as pumps.

The overall process efficiency (see Equation 2) sums the net thermal energy output of the process and the net power output and divides it by the total energy input of the fuel. The net thermal energy output is calculated by subtracting the thermal input to the process such as the energy contained in boiler feed water from the total thermal energy output of the process (thermal energy contained in products such as process steam and/or liquid/solid fuels).

For the purpose of this study, another definition of process efficiency referred to as the effective electric efficiency (see Equation 3) was used. This definition of efficiency was adopted from the United State Environmental Protection Agency (US EPA) (www.epa.gov) and it estimates the ability of a process to produce electric power after satisfying the energy needs of an end-user. It is the ratio of the net power output to the extra fuel required above what is needed to provide the end-user thermal energy.

Electrical efficiency:

$$\eta_{electrical} = \frac{E_{elec}\ power}{E_{th}\ biomass\ fuel} \tag{1}$$

Overall process efficiency:

$$\eta_{overall} = \frac{E_{elec}\ power + E_{th}\ process}{E_{th}\ biomass\ fuel} \tag{2}$$

Effective electric efficiency:

$$\eta_{electrical} = \frac{E_{elec}\ power}{E_{th}\ biomass\ fuel - E_{th}\ process/\alpha} \tag{3}$$

Where,

E_{elec} *power* = net electric power output (MW_{elec}).

E_{th} *biomass fuel* = thermal energy in the input bagasse feed (MW_{th}).

E_{th} *process* = net thermal energy output of process (MW_{th}).

α = the conversion efficiency of process technology used in place of the CHP to generate the net thermal output (80% efficiency based on the LHV of bagasse feed is assumed for the boilers used in this study [Bhatt and Rajkumar, 2001], hence $\alpha = 0.80$).

All energy efficiencies in this work were estimated based on the lower heating value (LHV) of the bagasse feed input. The LHV excludes the heat of vaporization of the water in the wet bagasse feed fuel and also does not account for the heat of condensation of water in the combustion flue gases since the water vapour component of the flue gas stream from the combustor is not condensed. The LHV of the bagasse feed at various moisture contents as used in this study were calculated from Equations 4 and 5 (Phyllis, 2005) below:

$$LHV_{dry} = HHV_{dry} - 2.442 \times 8.936 \times \frac{H}{100} \tag{4}$$

$$LHV_{ar} = LHV_{dry}\left(1 - \frac{x}{100}\right) - \frac{2.442x}{100} \tag{5}$$

Where,

HHV_{dry} = higher heating value of dry bagasse (18.65MJ/kg as used in this work; this is an average value estimated from the experimental results of Hugo, (2010)).

LHV_{dry} = lower heating value of dry bagasse.

LHV_{ar} = lower heating value of bagasse feed as received

x = moisture content of bagasse as received.

H = *weight* % hydrogen of dry bagasse

The heating values of bio-oil and biochar used in this work are 18.10MJ/kg and 24.65MJ/kg respectively and they are also average values from the work of Hugo (2010).

3.6 Validation of Models

All the combustion BMECP models are validated based on their steam production rates which are compared to industry standards. The rule of thumb in industry is to produce 2 kg of HP steam per kg of bagasse feed input (Pippo, 2009) at base conditions of 30bar pressure and 50% bagasse moisture content. Also it is known from literature that the overall efficiencies of CHP systems are in the range of 60% – 85% (www.epa.gov). The models are thus further validated by estimating their efficiencies and comparing them to this value. Aside overall process efficiency, the pyrolysis based models are also validated by comparing the yields of products from the various pyrolysis reactors to the yields of products obtained from the experimental work of Hugo (2010).

CHAPTER FOUR

Process Modelling

4.1 Introduction

This chapter provides a detail description of the various BMECP plants scenarios as evaluated in the Aspen Plus simulation software. This is then followed by the presentation and discussion of process model results. Process flow diagram descriptions are provided for both the combustion and pyrolysis BMECP model scenarios and are based on the overall process flow diagrams shown in Appendices A1 and A2, respectively.

4.2 Aspen Plus® Model of Combustion BMECP Plant

As stated in section 3.2.2 above, the combustion BMCEP plant was modelled based on the Rankine Cycle with modification for steam extraction. The overall process flow diagram is presented in appendix A1. For simplicity, the overall PFD was divided into "Areas" in Aspen Plus® and modelled accordingly. The process areas modelled are as follows:

- Area 1000 – Feed combustion
- Area 2000 – Heat recovery and Steam production
- Area 3000 – Steam Turbine and Power cycle
- Area 4000 – Utilities

The PFDs as well as the names of unit operation blocks and flow streams used in the description of the various subsections (Areas) are also provided in appendix A1. All combustion models developed have similar PFDs, and so to avoid repetition, the description of the various flow areas fits for all the combustion model scenarios simulated in Aspen Plus®.

The complete set of PFDs together with the mass and energy balance results of all flow sheet sections for all modelled scenarios can be found on the attached CD.

4.2.1 Area 1000 – Feed Combustion

The objective of this area is to produce hot combustion gases from the biomass feed. The main unit operation block in this area is the biomass combustor (see Figure 15). Though the biomass combustor forms an integral part of the boiler in real life application, it was modelled separately in this work. Bagasse feed stream (1001) at a given moisture content (depending on the scenario) and combustion air stream (1002) are fed to the biomass combustor QCB 1001, which was modelled as a stoichiometric reactor (RSTOIC) in Aspen Plus® and, all combustion reactions as well as the yield of combustion products are generated by the simulation software. The combustor was modelled to operate at adiabatic conditions and this was controlled by a design specification function in Aspen Plus®. This was done to satisfy the condition that all heat energy used in the production of steam is supplied only by the biomass feed.

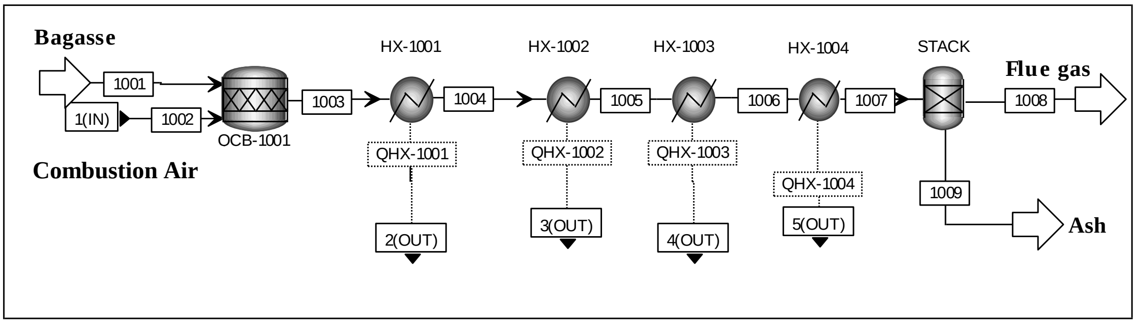

Figure 15: Combustion BMECP - Area 1000 PFD

To ensure complete combustion, air was supplied in excess of the stoichiometric amount at a preheat temperature of 250°C. Preheating was done to improve the efficiency of the combustion process and hence the efficiency of the steam boiler (EDUCOGEN and INESTENE, 2001). The percentage excess air employed in this study is in the range of 40 to 50% (value depending on the scenario under consideration). This ensures that the percentage mass of oxygen in the flue gas exiting the combustor stack is maintained above 6% in order to meet the minimum emission standards stipulated for combustion installations (Department of Environmental Affairs and Tourism, 2008). Heat for air pre-heating is provided by the combustion flue gases stream (1003) and the flow of air is controlled by a design-spec function AIRTEMP to ensure that the desired preheat temperature and the percentage excess amount is obtained. Heat is recovered from the combustor flue gases stream (1003) to generate High Pressure (HP) steam.

4.2.2 Area 2002 – Heat Recovery and Steam Production

In this area, heat is recovered from the combustor flue gases and is used to produce HP steam (stream 2008) in the boiler. For simplification and to ensure easy flow convergence, the boiler was modelled as a series of heat exchangers HX-2001, HX-2002, and HX-2003, which represents the steam side of the boiler. These were coupled to the heat exchangers HX -1001, HX-1002 and HX-1003, respectively, in area 1000, representing the process side of the boiler in Aspen Plus®. Boiler feed water stream (2004) is pumped from the feed water tank to the feed water preheater (HX-2001/HX-1001), where it exchanges heat with combustion flue gases. The preheated water is sent to the steam drum, which was modelled by the HX-2002/HX-1002 heat exchanger couple. Here saturated steam is produced and then fed to the superheater section of the boiler, modelled by the HX-2003/HX-1003 heat exchanger couplet to generate HP steam. Depending on the scenario that is being considered, the steam is raised to a certain degree of superheat (350-520 ^{o}C) using energy from the hot combustion gases. The amount of HP steam produced is limited by the superheat conditions required, hence a design-spec function WFLOW was set in Aspen Plus® to control the flow of boiler feed water such that the specified temperature and pressure of the HP steam required for a particular model is achieved.

The air supply for combustion (stream 2001) is also preheated to the required temperature of 250^{o}C in this area using part of the heat recovered from the flue gases in the HX-2004/HX-1004 heat exchanger couple (air preheater), prior to being sent to area 1000.

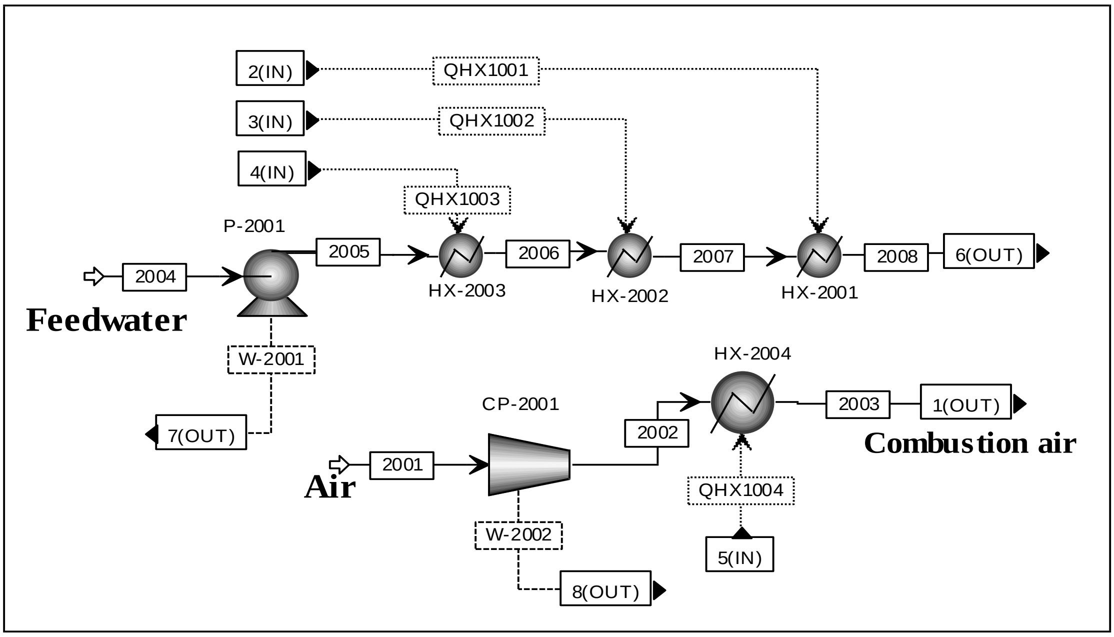

Figure 16: Combustion BMECP - Area 2000 PFD

4.2.3 Area 3003 – Steam Turbine and Power Cycle

Area 3003 simulates the electrical power production area of the combustion BMECP plant. Here HP steam (stream 3001) from the boiler is directed to a turbo alternator system/steam turbine (TB-3001, TB-3002) for electricity production and subsequent production of LP steam (stream 3004) for use by the sugar mill. All turbine systems used in this simulation were modelled as isentropic turbines working at an isentropic efficiency of 85% and a power loss of 2% was assumed for the power generation systems. Depending on the scenario, either a BPST or a CEST system is used to expand the steam for electricity production. The base case scenario uses a BPST system and in this case all HP steam coming from the boiler is expanded to a 2bar pressure LP steam and sent to the sugar mill for use as process steam. For scenarios using CEST system, HP steam at a certain superheat condition (superheat conditions higher than that used for BPST system) is first expanded in a high pressure turbine (TB-3001) to an intermediate pressure, which is then further expanded in an intermediate pressure turbine (TB-3002) to reduce the pressure of the steam to 2bar. The amount of LP steam required for the sugar mill (depending on the mill's steam efficiency) is extracted at this stage and the remaining steam expanded to condensing conditions of 0.2bar.

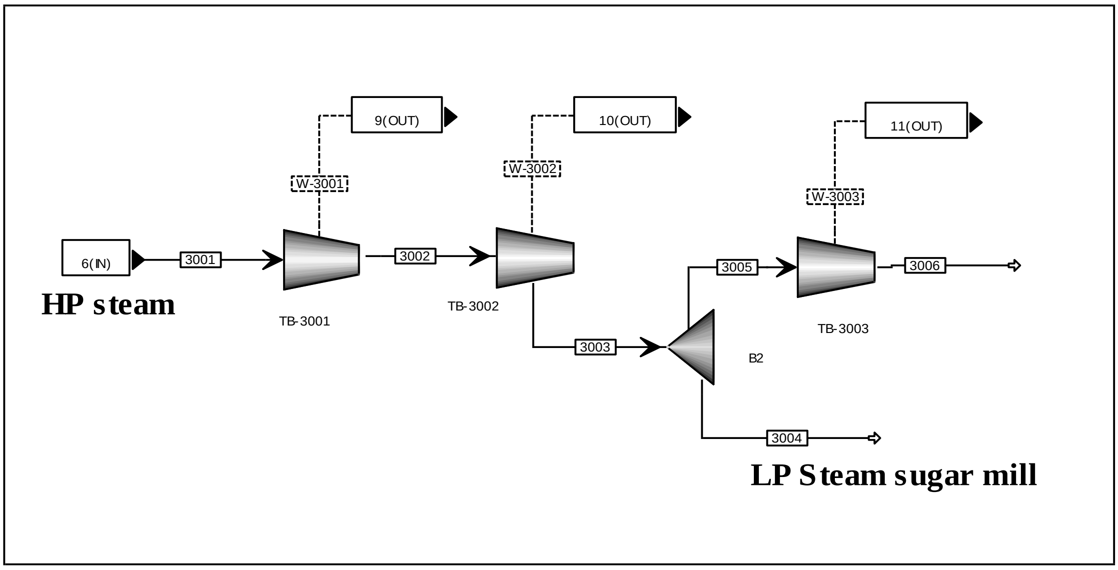

Figure 17a: Combustion BMECP - Area 3000 PFD (CEST system)

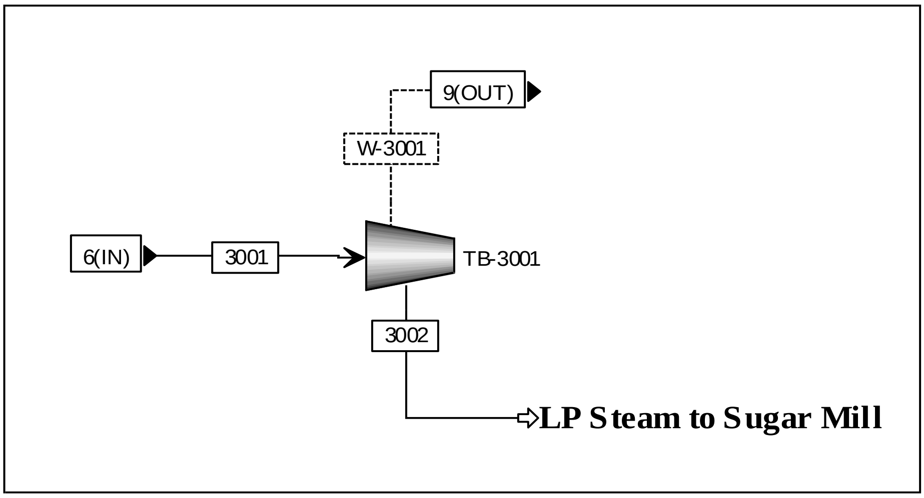

Figure 17b: Combustion BMECP - Area 3000 PFD (BPST system)

4.2.4 Area 4000 – Utilities

In the utilities area, all the electrical power produced by the BMECP plant as well those used in the process for running unit operation equipment, such as pumps and compressors, are combined to determine the net available power.

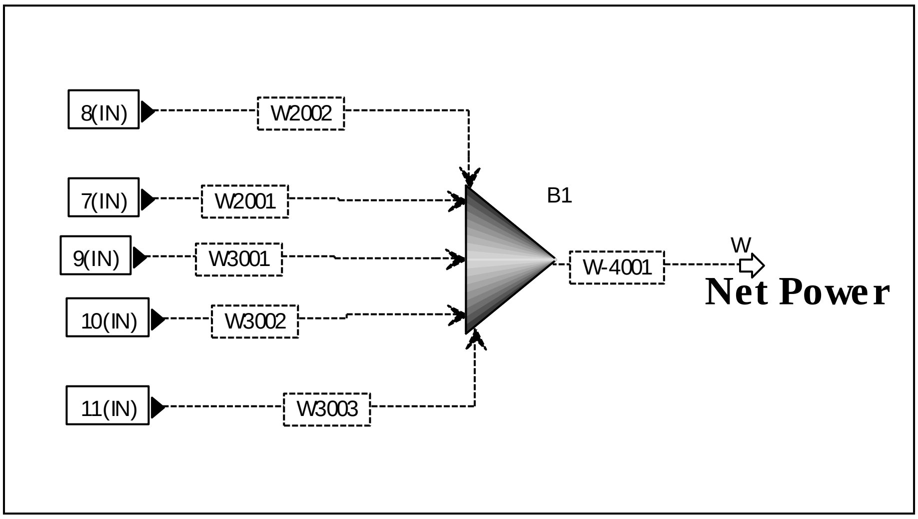

Figure 18: Combustion BMECP - Area 4000 PFD

4.3 Aspen Plus® Model of Pyrolysis-based BMECP Plant

The overall PFDs of the pyrolysis-based BMECP plants are presented in appendix A2 and A3. For simplicity, the PFDs were also grouped into various flow sheet sections as listed below:

- Area 1000 – Feed Pre-treatment
- Area 2000 – Fast Pyrolysis
- Area 3000 – Quench
- Area 4000 – Heat Recovery
- Area 5000 – Oil Recovery
- Area 6000 – Recycle
- Area 7000 – Steam and Power Cycle
- Area 8000 – Utilities

The PFDs of the various flow sheet sections listed above are also provided in appendix A2 and A3 and a complete set of each flow sheet section together with the mass and energy flows can be found on the attached *CD.*

Both the Pure Fast Pyrolysis BMECP and the Partial Fast Pyrolysis BMECP models developed were based on the method of Leibbrandt (2010) with some modifications and the data used was taken from the work of Hugo (2010). The modifications applied include the following:

- Changing of biomass feed composition to reflect the composition of sugarcane bagasse as obtained from the experimental work of Hugo (2010; see table 11). Although Leibbrandt's (2010) model was for sugarcane bagasse pyrolysis, the data used for biomass composition was that of wood, which was modified to reflect literature values as reported in the work of Piskorz et al. (1998).
- Modification of pyrolysis products yields from the pyrolysis reactor to agree with the yields obtained from experimental work of Hugo (2010).
- Inclusion of design specification functions to control the temperature of the pyrolysis reactor as well as the flow and generation of steam.
- Changing of the combustor model to work at adiabatic conditions. For the Partial Fast Pyrolysis BMECP scenario, the combustor is also modified to handle raw bagasse.

Description of the various flow sheet areas as modelled in Aspen Plus® are presented below. Because the Partial Fast Pyrolysis BMECP scenario is a variant of the Pure Fast Pyrolysis BMECP scenario, the PFDs for the flow sheet areas are much similar, hence the descriptions given below fit for both scenarios unless otherwise stated in which case the two different PFDs are both provided.

4.3.1 Area 1000 – Pre-treatment

The pre-treatment area consists of two main unit operations; drying and grinding of biomass. In this area, wet bagasse feed stream (1001) at given moisture content (50% or 48% or 46%) is dried to a lower moisture content (< 10%) using preheated air stream (1002), after which it is grinded to reduce its particle size. In the Partial Fast Pyrolysis BMECP, bagasse for the biomass combustor is first separated from the wet bagasse feed stream (1001) before the remaining is sent to the dryer (See Figure 19b). The feed dryer was modelled using a heater (DR-1001) and a flash (DR-1001FL) unit operation models in Aspen Plus® and its temperature was controlled such that the dried feed stream outlet (stream 1005) is at a moisture content below 10%, which is a necessary condition for fast pyrolysis. All moisture removed (stream 1004) is flashed off to the atmosphere and the dried bagasse fed to the pyrolysis section (Area 2000). The dryer air is preheated using heat recovered from condensing pyrolysis vapours (Area 3000). The flow and temperature of the air used for drying is controlled by a design-spec function DRYERAIR, which in turn controls the temperature inside the dryer to achieve the desired moisture content.

Grinding of the bagasse feed is important to reduce its particle size to 2mm or less as discussed in section 2.5.2. The bagasse grinder was not modelled in this study, however, the power required for grinding the feed was calculated according to the method of Mani et al. (2004), and added to the utilities area to establish the net power output of the process.

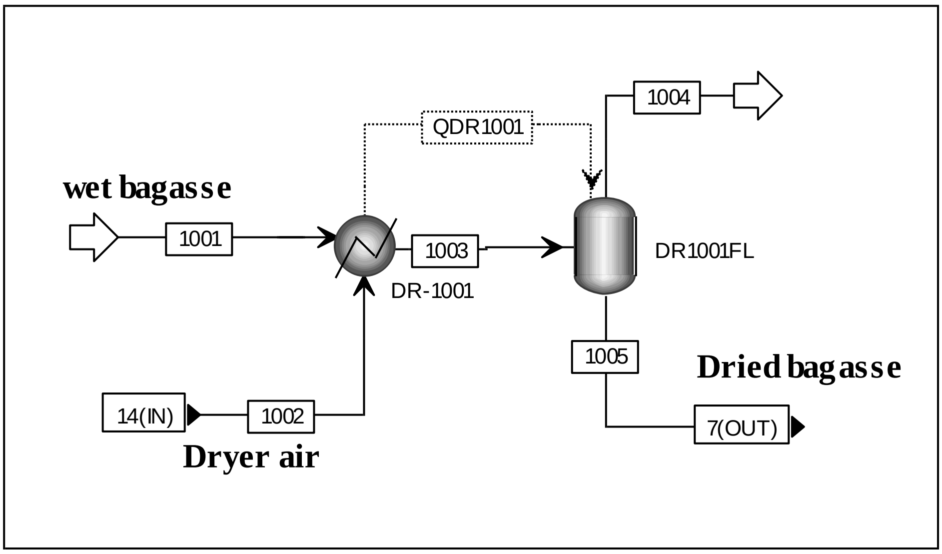

Figure 19a: Pure Fast Pyrolysis BMECP - Area 1000 PFD

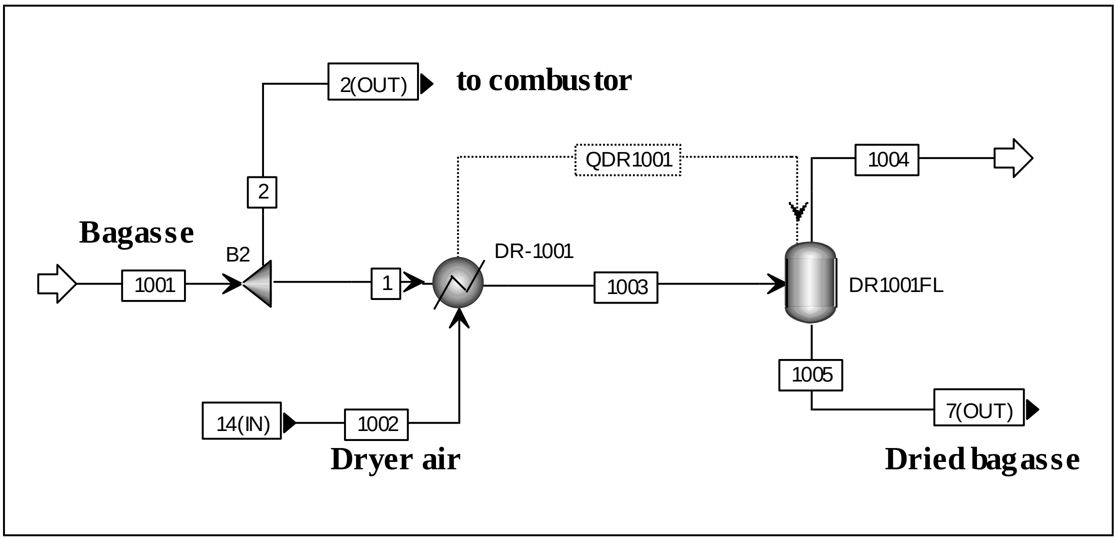

Figure 19b: Partial Fast Pyrolysis BMECP - Area 1000 PFD

4.3.2 Area 2000 – Fast Pyrolysis

The main equipment in this area is the fast pyrolysis reactor. A fluidized bed reactor was selected as the preferred choice for this work because of its versatile application and also due to the fact that data used in modelling the pyrolysis reactor was that obtained from experimental work carried out using a fluidized bed reactor (Hugo, 2010).

In this area, dried bagasse feed stream (1005) is mixed with recycled pyrolysis non-condensable gasses (stream 6008) and fed to the pyrolysis reactor (PY-2001), where the feed undergoes pyrolysis. The gas stream acts as a fluidizing medium (Note that in reality, an inert gas stream would be required to start off the process before switching over to the recycled pyrolysis non-condensable gasses). Heat for pyrolysis is supplied by the combustor (refer to sections 4.3.4 and 4.3.6) and the temperature of the mixture is raised to 500^{o}C (pyrolysis temperature) inside the reactor and is controlled by a design-spec function in Aspen, which ensures that a constant temperature is maintained within the reactor. The biochar/biomass combustor provides the heat required for pyrolysis.

A yield reactor models the pyrolysis reactor (PY-2001) in Aspen Plus® and the yields of products from the pyrolysis reaction were specified using measured experimental data from pyrolysis experiments previously conducted by Stellenbosch University (Hugo, 2010). After pyrolysis, the product stream from the reactor (2003), consisting of condensable and non-condensable biocrude components and biochar, is first sent to a cyclone (CY-2001), where biochar (stream 2005) is separated out. Biocrude components (stream 2004) are also sent to the quench (Area 3000) for bio-oil recovery. For Pure Fast Pyrolysis BMECP the biochar stream (2005) is sent to the combustor where it is combusted together with reactor off-gas to supply the heat for the process as shown in Figure 20a. The biochar stream in the Partial Fast Pyrolysis BMECP is stored as final product (see Figure 20b).

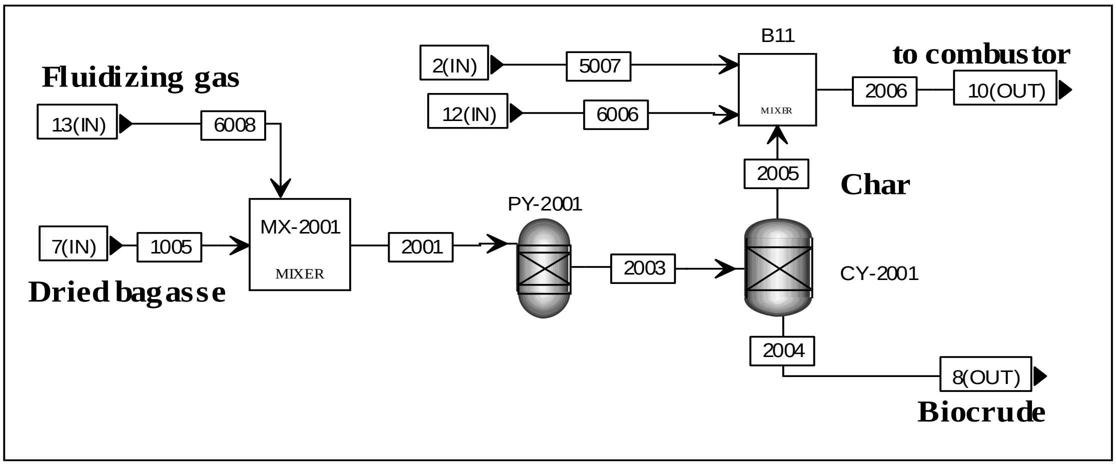

Figure 20a: Pure Fast Pyrolysis BMECP - Area 2000 PFD

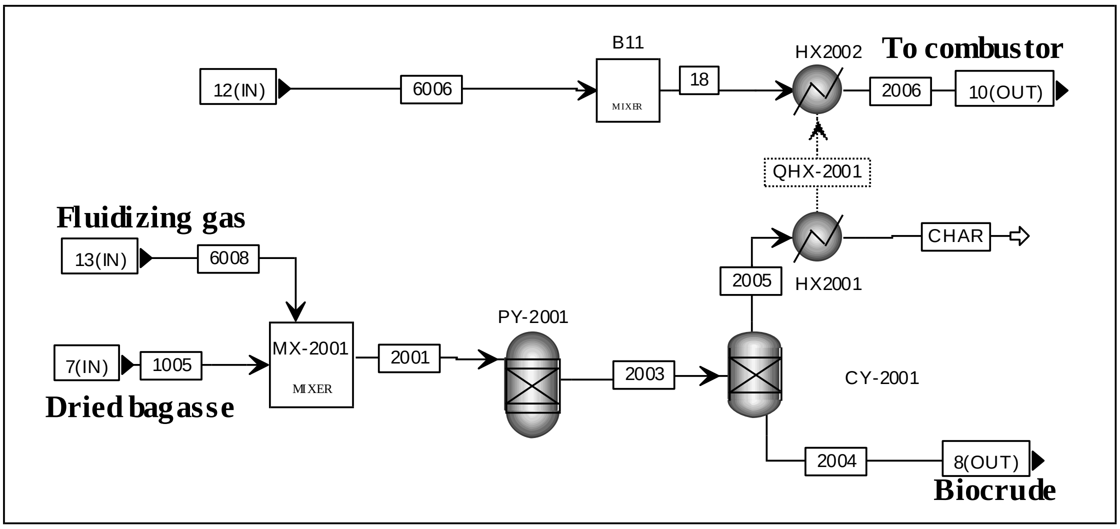

Figure 20b: Partial Fast Pyrolysis BMECP - Area 2000 PFD

4.3.3 Area 3000 – Quench

The quench area is where biocrude vapours are rapidly condensed using condensers (HX-3001+ and HX-3002+) arranged in series. Water is used as the condensing medium in the first condenser (HX-3001+) and the recovered heat is used to produce steam (stream 3010), which is then sent to the steam drum in the steam and power cycle section (A 7000). Air (stream 3006) is used in the second condenser for cooling and the heated air is what is used to dry the wet bagasse feed in the feed dryer. The flow of air is controlled using a design specification DRYERAIR in order to achieve the required temperature for drying. A scrubber (SC-3001) and an electrostatic precipitator (ESP-3001) are used to remove any remaining aerosols from the non-condensable gas vapours, after which all liquid recovered (streams 3004 and 3021) are sent to product recovery (AREA 5000) for storage, while the vapour stream (3020) exiting from the electrostatic precipitator is sent to recycle section (AREA 6000) and recycled for use as pyrolysis fluidising medium.

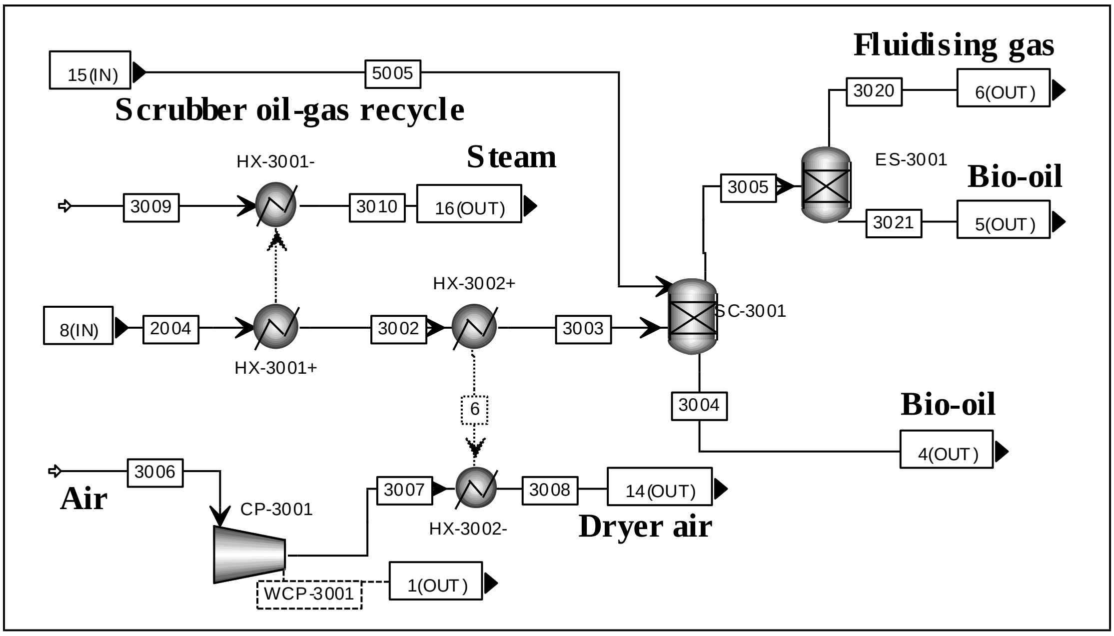

Figure 21: Pure/Partial Fast Pyrolysis BMECPs - Area 3000 PFD

4.3.4. Area 4000 – Heat Recovery

This area is where the heat energy required to run the pyrolysis reactor and to generate steam for subsequent electricity production is produced. A fraction of cleaned vapours from the recycle section (Area 6000) is mixed with a fraction of bio-oil and biochar and/or bagasse (depending on whether it is Pure or Partial Fast Pyrolysis scenario) and fed as fuel stream (10) to the combustor (QCB-4001) for combustion to generate heat energy in the form of hot flue gases. The combustion reactor was modelled as an adiabatic reactor and air (stream (4010)) at 40% excess above the stoichiometric amount was supplied to ensure complete combustion. The RSTOIC reactor unit operation model was used to model the reactor and Aspen Plus® calculates the appropriate combustion reactions. Heat is recovered from the combustor flue gases (stream 4001) and used to supply the heat for pyrolysis and to raise HP steam for electricity generation after which LP steam is recovered and sent to the sugar mill.

The flow of air (stream 4010) to the combustor is controlled by the design-spec function AIRCOMB, which ensures that the combustion is complete and that the fraction of oxygen in flue gas meets environmental standards (see section 4.2.1).

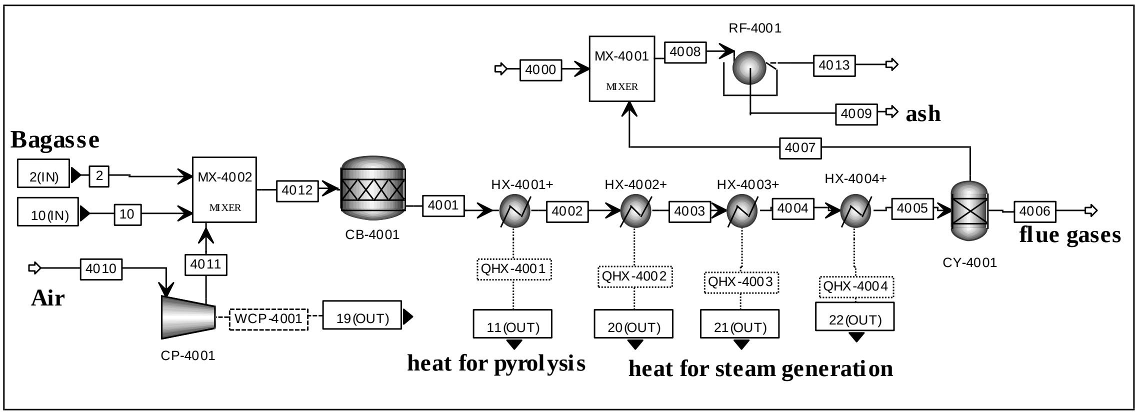

Figure 22a: Partial Fast Pyrolysis BMECP - Area 4000 PFD

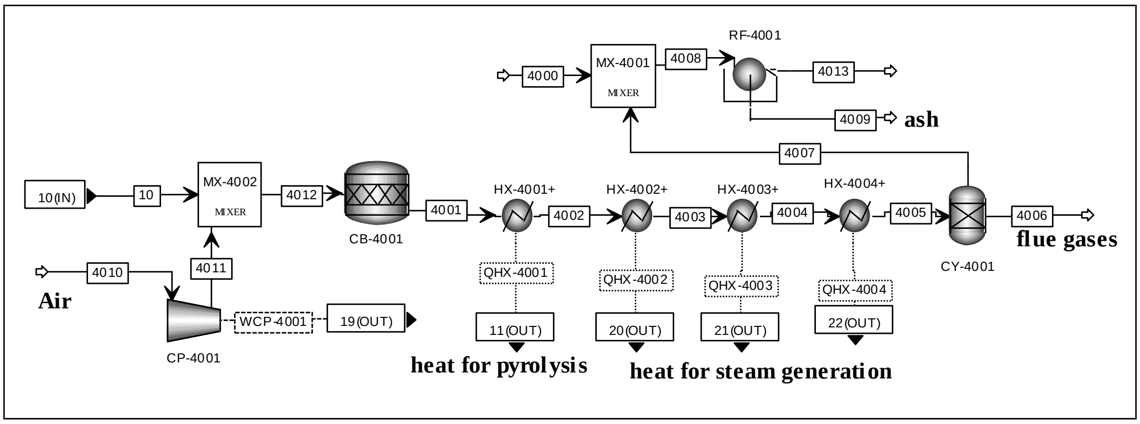

Figure 22b: Pure Fast Pyrolysis BMECP - Area 4000 PFD

4.3.5 Area 5000 – Oil Recovery

Bio-oil product streams (3004, 3021 and 6003) from the AREA 3000 and AREA 6000, are mixed together in the mixer (T-5001) and the resulting mixture is pumped to a cooler (HX-5001+) for cooling and then sent to product storage. Before storage, a small fraction of the oil is separated (see Figure 23a and Figure 23b) and sent to AREA 3000 for use as recycle stream oil (stream 5005) in the scrubber to increase the recovery oil. For the Pure Fast Pyrolysis BMECP some oil is also separated and sent to the combustor (see Figure 23b) where it is combusted together with biochar and recycled pyrolysis gases to provide the energy for pyrolysis and steam generation.

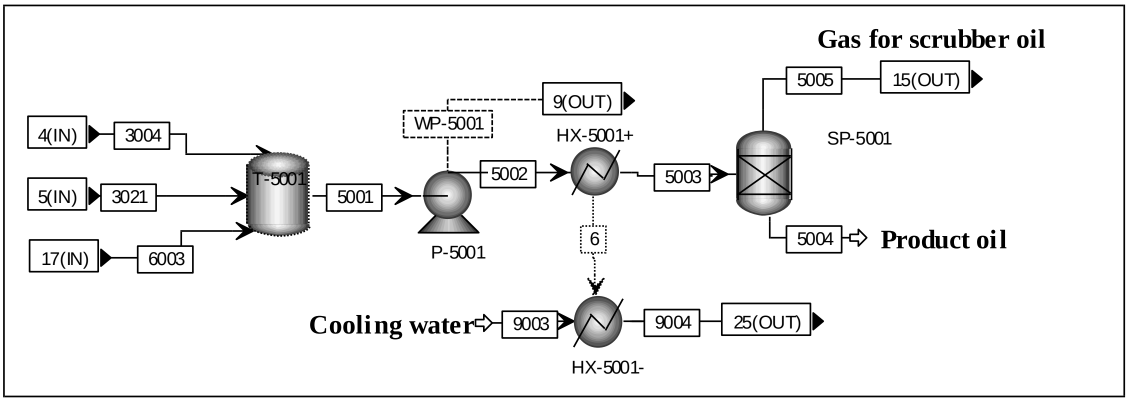

Figure 23a: Partial Fast Pyrolysis BMECP - Area 5000 PFD

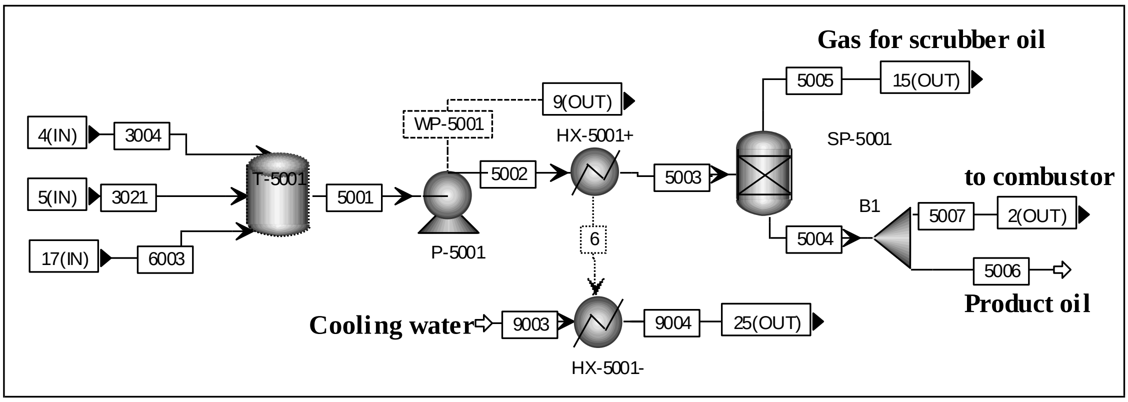

Figure 23b: Pure Fast Pyrolysis BMECP - Area 5000 PFD

4.3.6 Area 6000 – Recycle

In the recycle area, vapours (stream 6001) from the electrostatic precipitator in AREA 3000 are further cooled to condensing temperature in a condenser (HX-6001) using a chilled water stream (9006). Product from the condenser is sent to a flash drum (FL-6001) to recover additional oil product (stream 6003), which is separated out and sent to AREA 5000, where it is added to oil product from AREA 3000 for storage. Vapour stream (6004) from the flash drum is compressed and a part of it recycled for use as fluidizing air (stream 6007) in the pyrolysis reactor and the rest (stream 6006) sent to the combustor for process heat generation. For convergence purposes, the heat for pyrolysis QHX-4001 was added to the fluidizing air in this area before sending it to the pyrolysis reactor. Heat added was controlled using a design spec function FAIRTEMP to ensure that the inside temperature of the pyrolysis reactor is maintained at 500^{o}C. Figure 24 shows the PFD for this area and is common for both fast pyrolysis-based BMECP models.

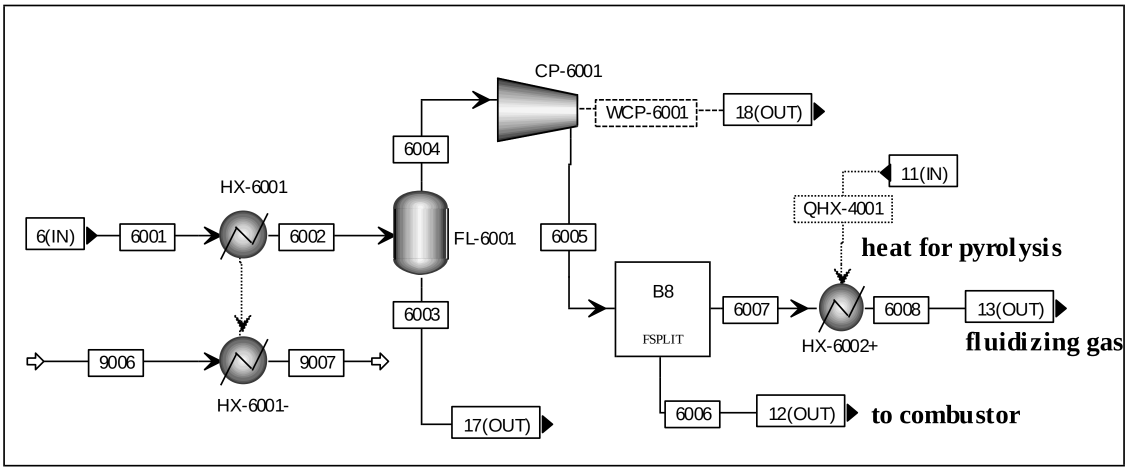

Figure 24: Pure/Partial Fast Pyrolysis BMECP - Area 6000 PFD

4.3.7 Area 7000 – Steam and Power Cycle

The steam cycle consist of the steam boiler (modelled as series of heat exchangers, HX-7001, HX-7002, and HX-7003 coupled to HX- 4001, HX-4002 and HX-4003 respectively) and the steam turbine (TB 7001). The method used in modelling the steam boiler is very similar to that used to model the steam boiler for the combustion BMECP plant; a thorough description is given in section 4.2.2 and is thus not repeated here.

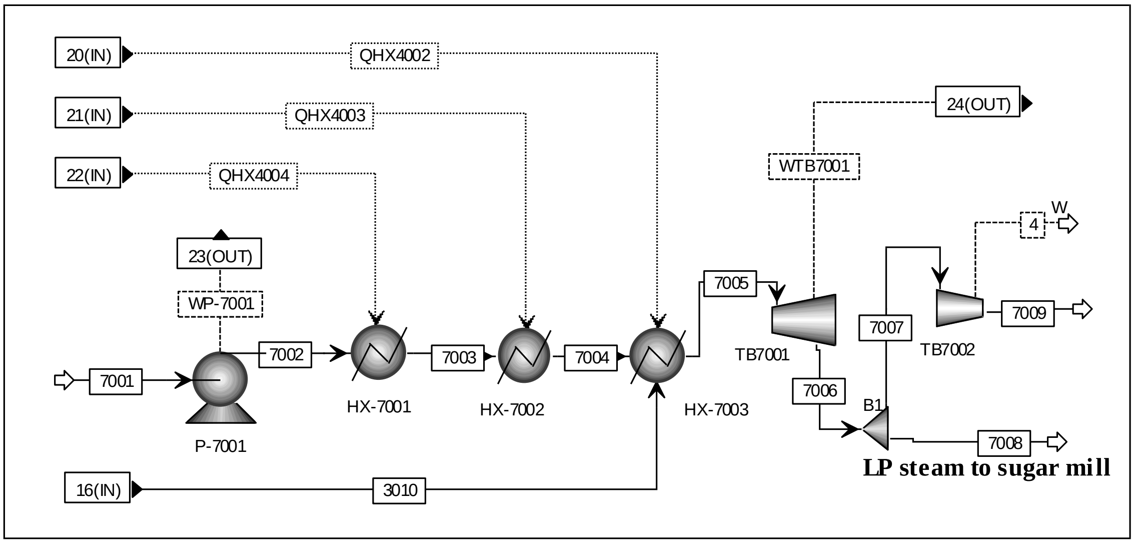

Figure 25: Pure/Fast Pyrolysis BMECP - Area 7000 PFD

In this area, heat from the combustor is used to raise HP steam which is then combined with the steam (stream 3010) produced from the first condenser in AREA 3000 and then sent as stream (7005) to the steam turbine for electricity generation and the production of LP steam (stream 7011) for the sugar mill. The turbine operates at an isentropic efficiency of 85% and a 2% loss in electricity was assumed.

The flow rate of boiler feed water (stream 7001) was controlled by a design-spec function BFWFLOW to ensure that the specified superheat conditions of the HP steam are achieved.

4.3.8 Area 8000 – Utilities

This section (Figure 26) of the flow sheet combines all the electrical power produced by the pyrolysis BMECP plant as well as those consumed by process equipment such as the biomass grinder, pumps and compressors to calculate the net electrical power. Process water utilised in the process is also included in the utilities section.

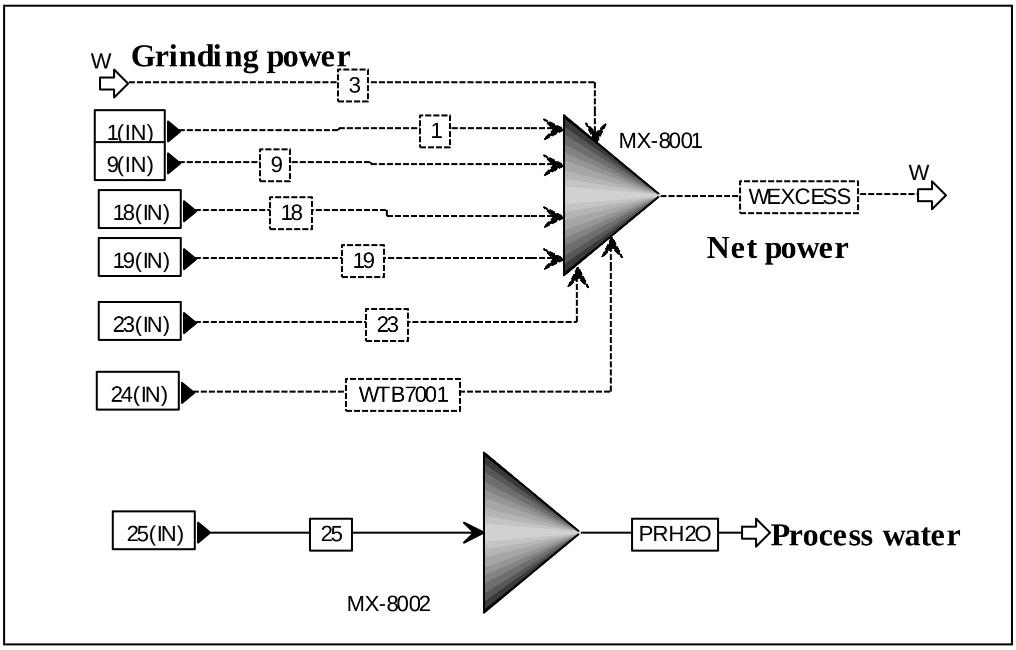

Figure 26: Pure /Partial Fast Pyrolysis BMECP - Area 8000 PFD

4.4 Process Simulation Results and Discussions

The results of the various process simulations developed are presented and discussed below. All process models are developed in order to meet the energy needs (see Table 11 below) of two sugar mill conditions, that is a 50% steam on cane mill and a 40% steam on cane mill (herein referred to as 'less efficient mill' and 'efficient mill' respectively) based on a 300 ton cane/h throughput mill (81 ton wet bagasse @ 270 kg wet bagasse/ton of cane crushed (tch)). The details of the mass and energy balance generated from Aspen Plus® for the various flowsheet sections are given in Appendix A.

Table 11: Hourly process energy consumption of two sugar mills

Process Energy	Type of Sugar Mill	
	Less efficient mill	Efficient mill
Steam @ 2bar &120°C, tons/h	150	120
Electricity, MW	6	6
kW/tch	20	20

4.4.1 Combustion BMECP Results

4.4.1.1 Steam Production

By using all available bagasse (81 tons of wet bagasse/h) as fuel input to the combustion BMECP, the quantity of HP steam generated for different boiler pressures and bagasse moisture contents is shown in Figure 27 below.

From Figure 27, it is seen that the quantity of steam produced at each given boiler pressure and bagasse moisture content exceeds the steam requirements of both the less efficient and efficient sugar mills. At the current cogeneration operating conditions (30 bar pressure and 50% bagasse moisture content) in the sugar industry, the rule of thumb is to produce 2 kg of HP steam per kg of bagasse feed input (Pippo, 2009). From the process simulation, 161.73 tons of HP steam is produced at this base case condition as can be seen in Figure 27, thus validating the credibility of the simulated model. Models were further validated based on criteria such as product yields from reactors and process efficiencies which are discussed in further sections.

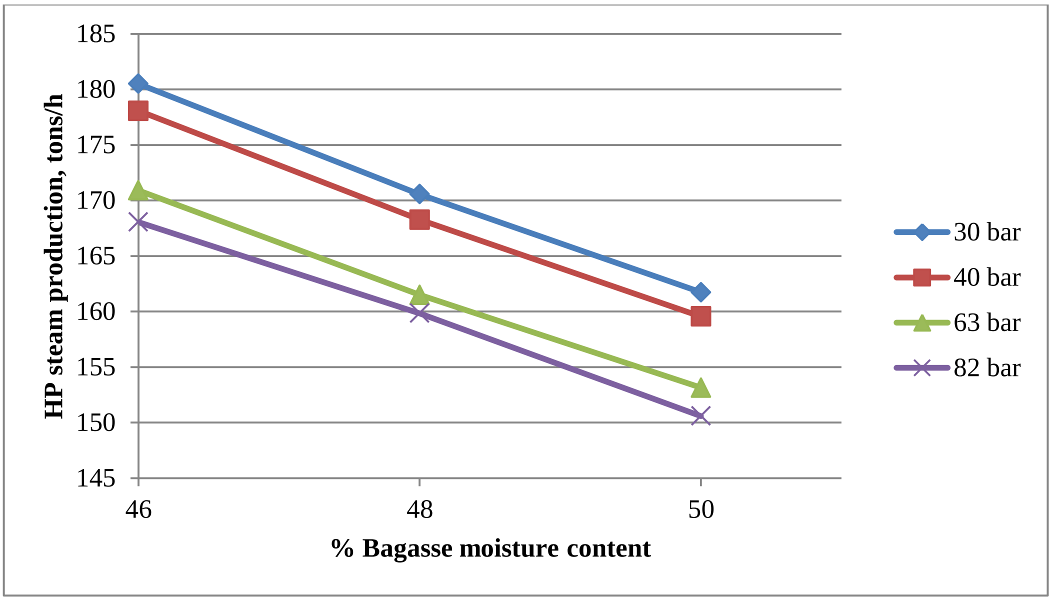

Figure 27: HP steam generation capacity of a combustion BMECP at varying pressures and bagasse moisture contents (based on a 300 ton cane/h throughput)

The effect of bagasse moisture content on HP steam production is depicted in Figure 27. It is shown that the production of HP steam increases with decreasing bagasse moisture content at a given boiler pressure. For instance at 30 bar pressure, there is an increase in HP steam generation from 161.73 ton at 50% bagasse moisture to 180.50 tons at 46% bagasse moisture content, representing 11.6% increase in steam generation for a 4% drop in fuel moisture content. Similar percentage increase in steam production is recorded for higher boiler pressures. This trend in steam production is as a result of the increase in the heating value of the bagasse fuel with decreasing moisture content as depicted in Equation 5 (See section 3.4). The increase in heating value together with a decrease in heat of vaporisation of water from the bagasse (also decreasing with decreasing bagasse moisture), increases the adiabatic combustion temperature within the combustion chamber of the boiler thereby making more heat available for steam generation.

As also seen from Figure 27, there is an inverse relationship between HP steam production and operating pressure of the steam boiler. For instance there are decreases of 1.34%, 5.31% and 6.88% in steam production as boiler pressure increases from 30bar pressure to 40bar, 63bar and 82bar respectively at a given moisture content of bagasse feed. The quantity of heat available from the combustor for steam generation is the same for a given bagasse moisture content since the same feed throughput is fed to the combustor. However, an increase in the boiler pressure requires an increase in the degree of superheat of the HP steam hence some portion of the available heat is used to achieve the required superheat conditions leading to a corresponding decrease in the quantity of HP steam generated.

4.4.1.2 Electricity production

The total net electricity output (MW) of the combustion BMECP models are shown in Figure 28a for the less efficient mill. Figure 28b shows the specific electricity production achievable as well the corresponding surplus electricity after meeting the demands of the sugar mill. Figures 28a and 28b show that regardless of the type of cogeneration system used (BPST for the 30 bar base case or CEST for the higher steam pressures), the net production of electricity as well as the specific electricity and surplus electricity increases as the moisture content of the bagasse feed decreases. This is as a result of the increase in HP steam production with decreasing moisture content as explained in section 4.4.1.1. An increase in the quantity of steam increases the mass flow rate of HP steam supplied to the turbine and hence an increase in the electricity generated.

The effect of pressure on electricity production is also seen in Figures 28a and 28b. The graphs show a direct relationship between pressure and electricity production. It is seen from figure 28a that at given moisture content of bagasse supplied, the production of electricity using a BPST system operating at 30 bar steam pressure is less than that of the CEST systems operating at the higher pressures of 40bar, 63bar and 82bar. This is caused by the low conversion efficiency of the BPST system. One interesting finding is that although there is a reduction in the generation of HP steam with increasing boiler pressure as shown in Figure 27, the resulting power obtained from the CEST systems at any given moisture content shows an opposite relation. This is caused by the increase in the efficiency of conversion of heat energy to electrical energy since an increase in boiler pressure increases the maximum temperature of the steam–turbine cycle (Mbohwa, 2003). Refer to Appendix C for the relationship between efficiency and temperature.

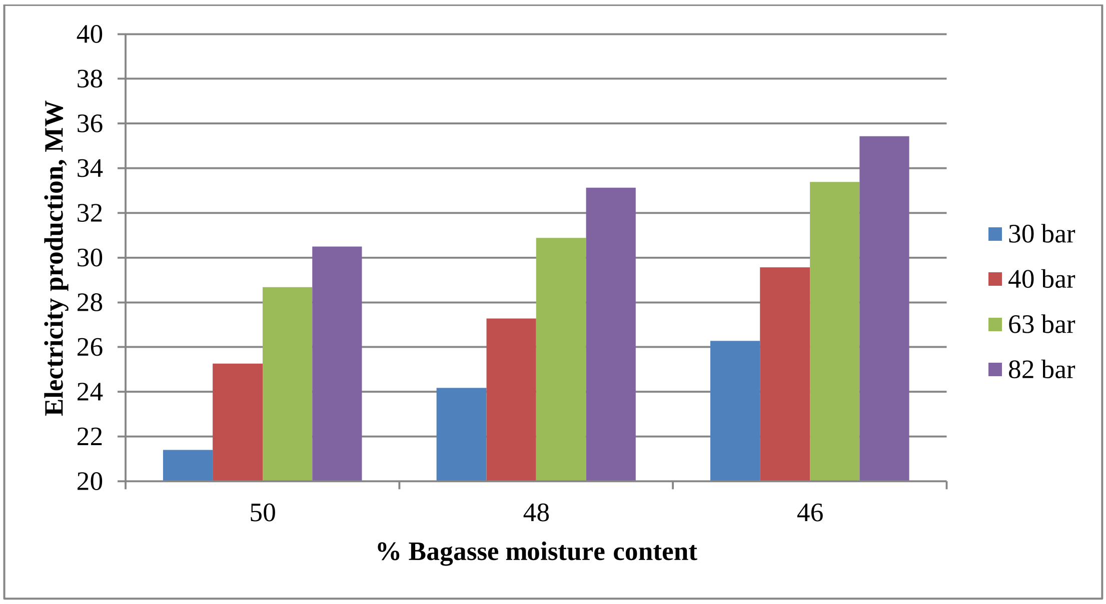

Figure 28a: Total net electricity output of combustion BMECP for less efficient mill

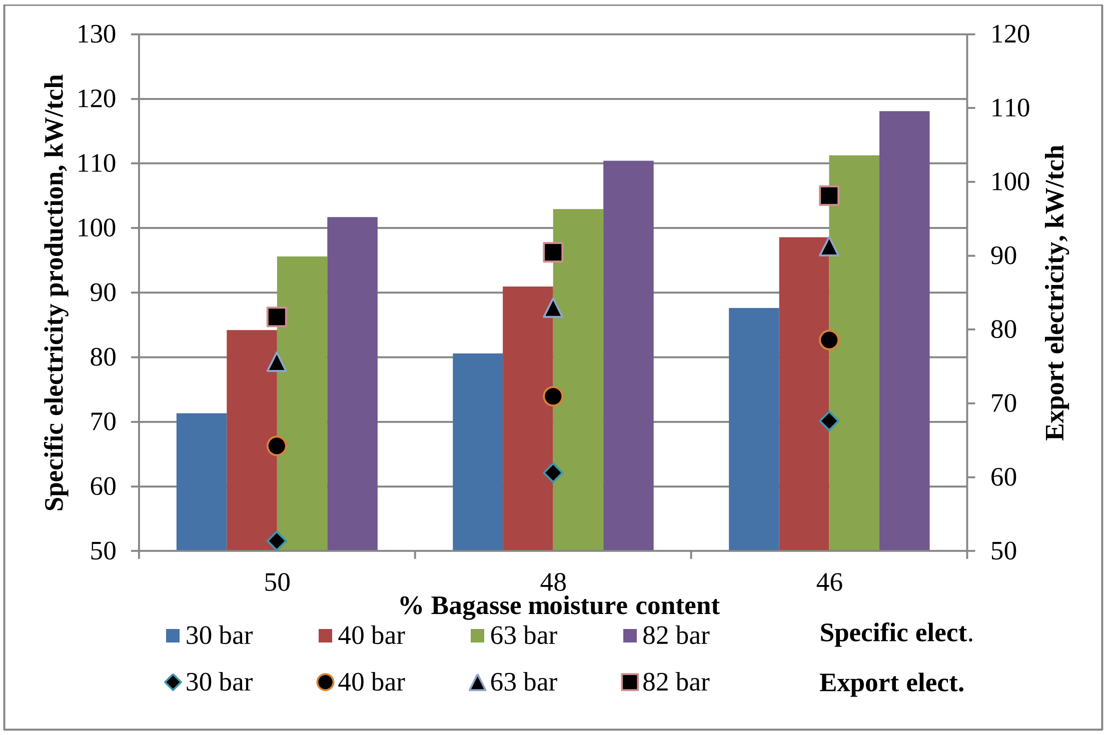

Figure 28b: Specific and Export electricity of combustion BMECP for less efficient mill

The trend in electricity generation is also observed for the case of the combustion BMECP modelled for the efficient mill scenario as shown in Figure 29a and Figure 29b. However, a comparison of Figures 29a and 29b to Figures 28a and 28b shows that at any given pressure and bagasse moisture content, the power produced in the case of the efficient mill is always greater than that of the less efficient mill. This increase in electricity production is directly related to the thermal energy/steam demand of the sugar mill. At lower % steam on cane, more steam is made available for further expansion in the steam turbine to condensing temperatures after the extraction of LP steam to the sugar mill. This results in the generation of more electricity over and above that obtained for a mill operating on a higher % steam on cane. As shown above in Table 11, the less efficient mill requires 150 tons/h of LP steam, while the efficient mill requires 120 tons/h of LP steam, giving an indication that the quantity of steam available for further generation of electricity after steam extraction is greater in the case of the efficient mill than the less efficient mill, hence the increase in electricity generated.

The results show that the sugar mill can produce as much as 105 kW/tch (see Figure 29b) of surplus electricity for sale by improving on the steam efficiency of the sugar making process as well as investing in advanced cogeneration systems operating at higher boiler pressure.

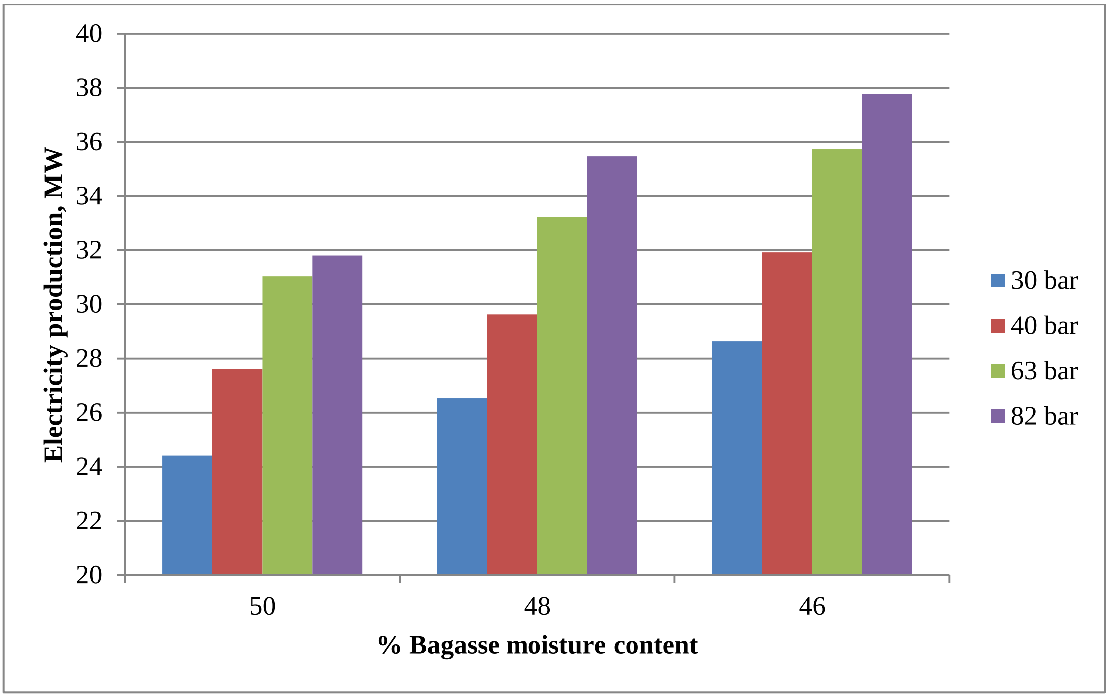

Figure 29a: Total net electricity output of combustion BMECP for efficient mill

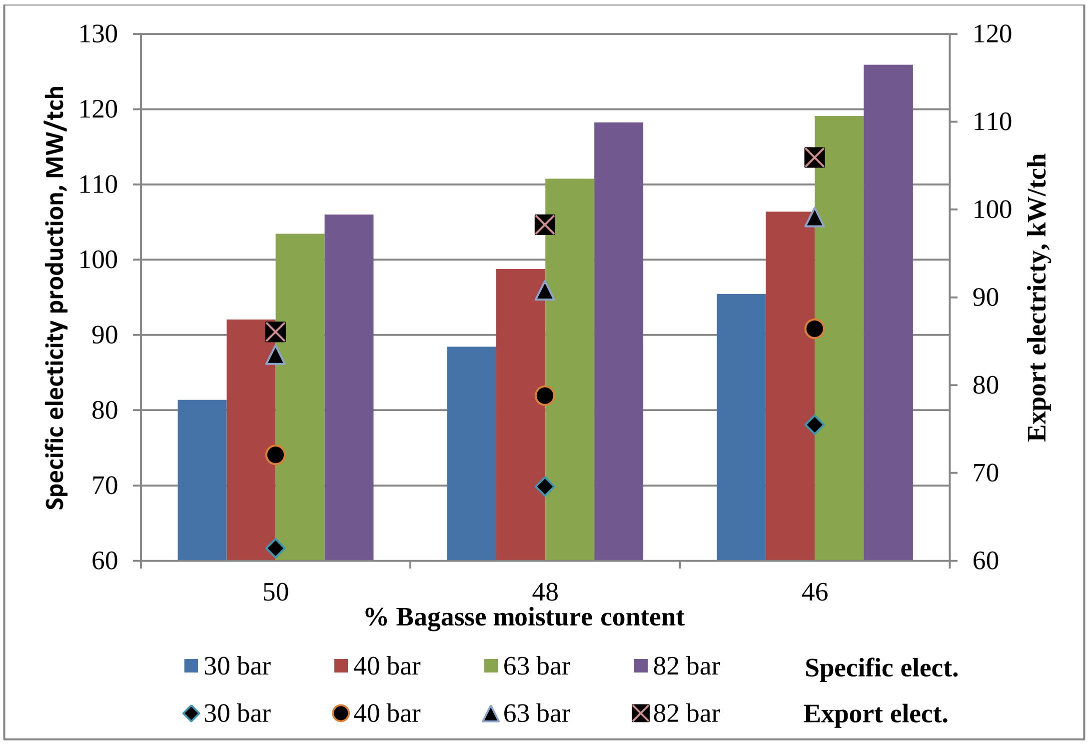

Figure 29b: Specific and Export electricity of combustion BMECP for efficient mill

4. 4.1.3 Energy Efficiency

The electrical efficiencies of the combustion BMECP's at varying cogeneration pressures and bagasse moisture contents are shown in Figure 30 for the less efficient mill and in Figure 31 for the efficient mill. From both figures, it can be seen that the electrical efficiency of the BMECP system increases with increasing pressure for each given bagasse moisture content. This is as a result of the increase in superheat conditions of the HP steam going to the steam turbine, which subsequently increases the maximum cycle temperature of the system, leading to an increase in the conversion efficiency of thermal energy of steam to electrical energy. Improvement in the cogeneration system from a BPST (30bar HP steam pressure) to CEST system results in a significant increase in electrical efficiency. From Figure 30 changing the turbo alternator system from a BPST system to CEST system pushes the electrical efficiency upward from 12.82% to 18.27% for an 82bar CEST turbo alternator at a bagasse moisture content of 50%. Also an increase in electrical efficiency from 14.63% (BPST) to19.10% (CEST) is obtained for the efficient mill as seen in Figure 31. Similar trends of electrical efficiency increase with pressure are recorded for the other bagasse moisture contents of 46% and 48% in both instances of the sugar mill (refer to Figures 30 and 31).

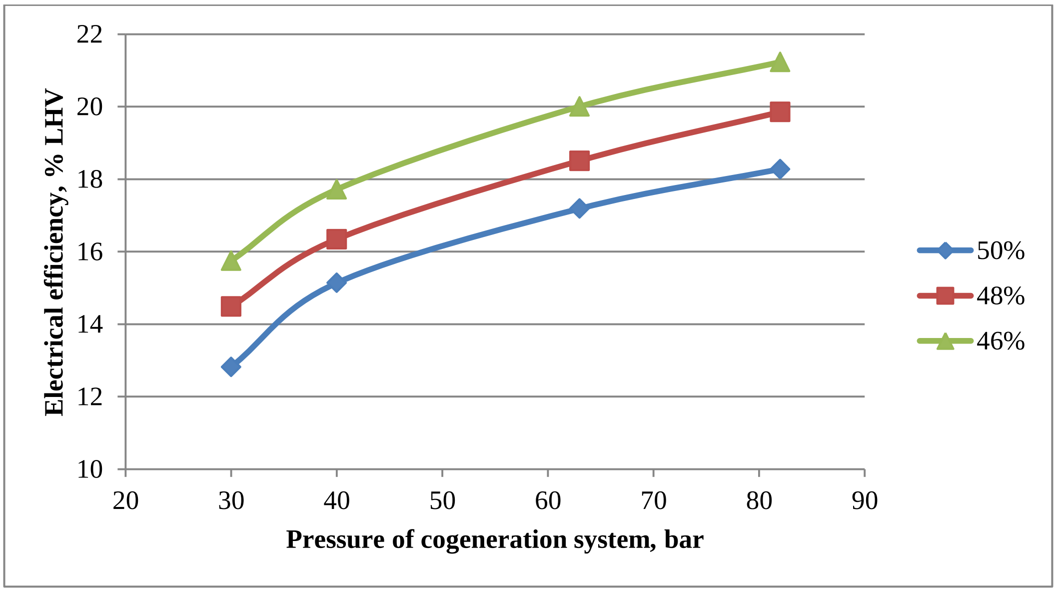

Figure 30: Electrical efficiencies of combustion BMECP for less efficient mill

From Figures 30 and 31, it is shown that the electrical efficiency also increases with decreasing moisture content of bagasse feed fed to the combustor. As explained earlier, a decrease in bagasse moisture content results in the increase in steam generation and hence an increase in the amount of electricity which in turn increases the electrical efficiency. By comparing Figure 30 and Figure 31, the effect of sugar mill efficiency is again made evident.

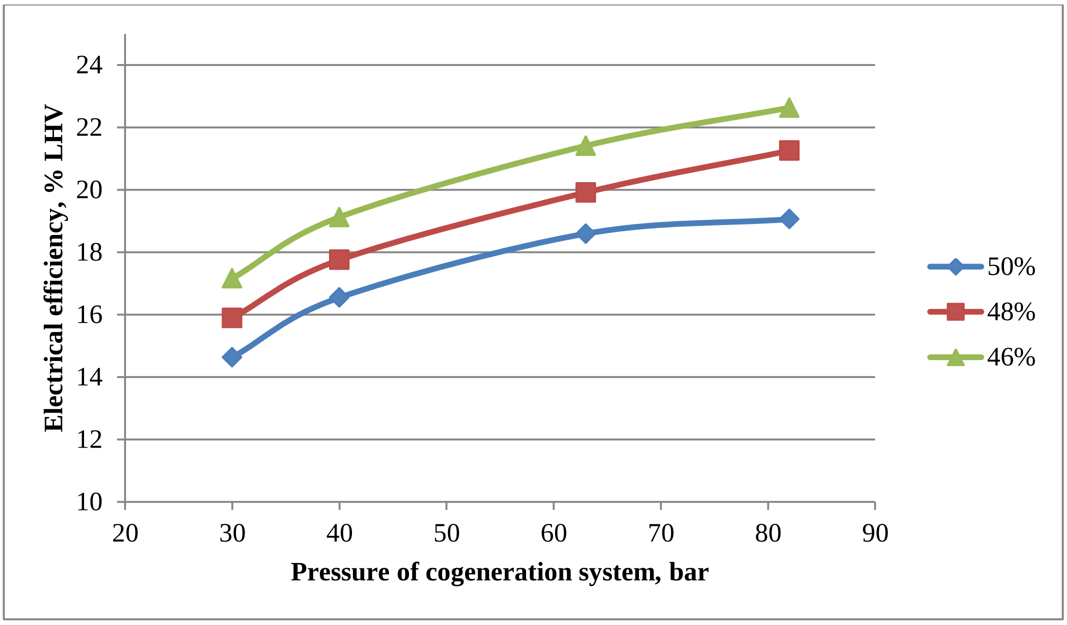

Figure 31: Electrical efficiencies of combustion BMECP for efficient mill

The overall system efficiencies of the various combustion BMECP models are also shown in Figure 32 and Figure 33 below for the less efficient and the efficient sugar mills respectively.

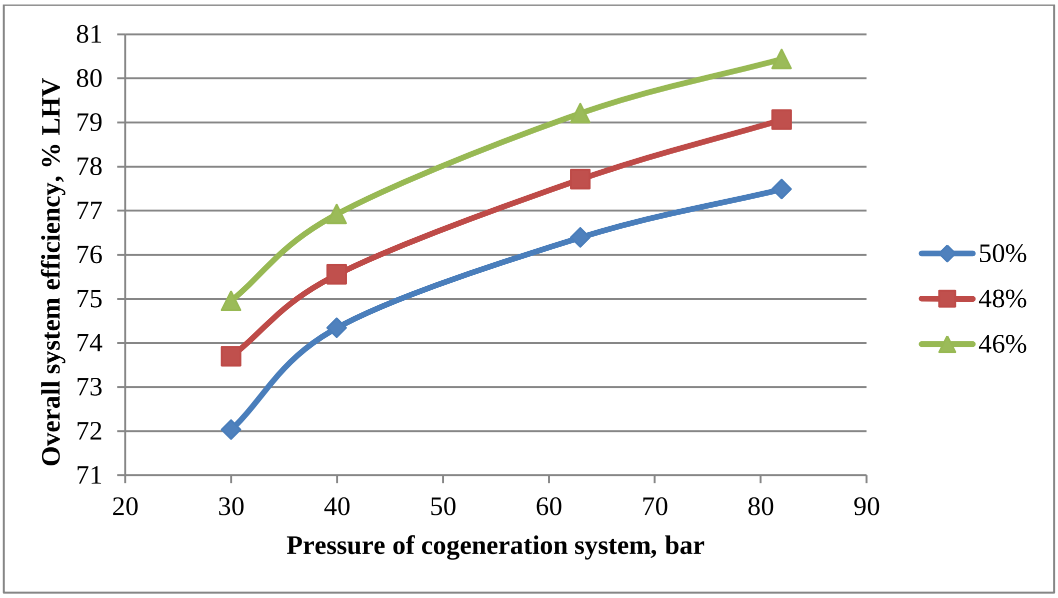

Figure 32: Overall system efficiencies of combustion BMECP for less efficient mill

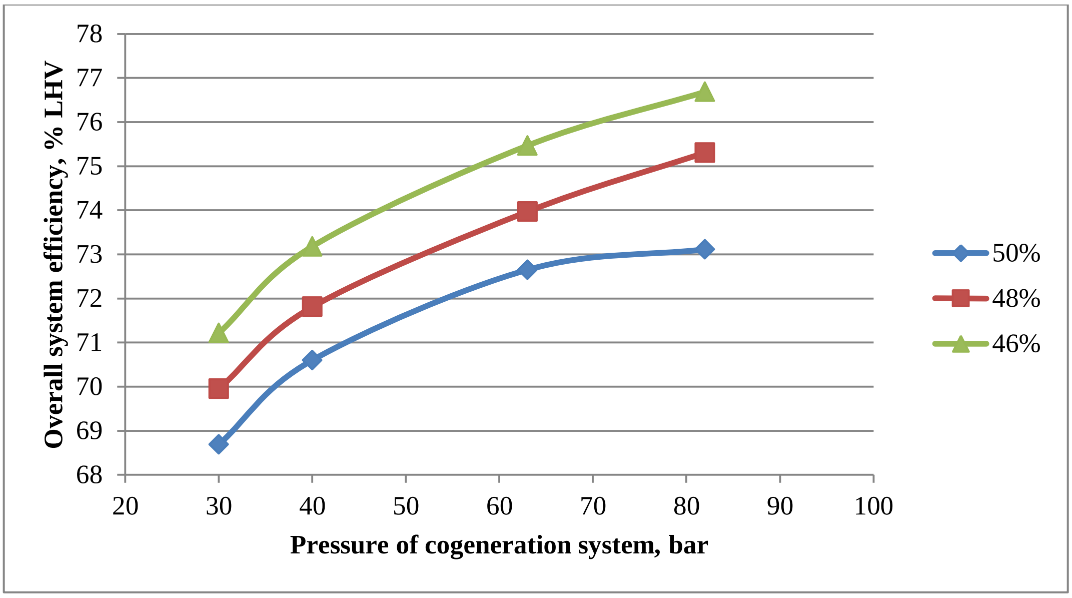

Figure 33: Overall system efficiencies of combustion BMECP for efficient mill

Similar to the electrical efficiencies, it can be seen from both figures that the efficiencies of the overall systems increase with increasing cogeneration system pressure and decreasing moisture content of bagasse feed in both instances of the sugar mill. The overall system efficiencies of typical CHP systems are in the range of 60% - 85% (www.epa.gov). By comparison, it can be noticed from both Figure 32 and 33 that the overall system efficiencies as obtained from the combustion BMECP models in this study fall within this range. The lowest and the highest overall system efficiencies obtained are 72.03% and 80.44% respectively for the less efficient mill (see Figure 32) and 68.69% and 76.69% respectively for the efficient sugar mill (see Figure 33).

Although the electrical efficiencies of the combustion BMECP models for the efficient mill are higher than those of the less efficient mill at any given bagasse moisture content and cogeneration system pressure, a comparison between Figure 32 and Figure 33 shows the opposite for overall system efficiency. The high thermal output from the various combustion based BMECP plants to the less efficient mill (150 tons/h of LP steam) accounts for the reason why their overall system efficiencies are higher than those of the efficient mill which requires a low thermal output (120 tons/h of LP steam) from the BMECP plants for its operation.

4.4.1.4 CO_2 Savings/Environmental Impact

Table 12 shows the savings in CO_2 made by using combustion BMECP to convert sugar mill biomass into electricity/energy. Results are shown for both the efficient mill case and the less efficient mill case. From Table 12, it is seen that CO_2 savings for the efficient mill at any given boiler pressure and biomass moisture are higher than those of the less efficient mill mainly because of the higher values of electricity generated by the combustion BMECP under this mill condition. The lower energy requirement of this mill (efficient mill) makes available more surplus bagasse for subsequent conversion into electricity.

Table 12: Estimated CO_2 Savings of Combustion BMECP for an Efficient and Less efficient Sugar Mills

% Bagasse moisture	CO_2 Savings, tons/h							
	Less Efficient Mill				***Efficient Mill***			
	BPST	**CEST**			**BPST**	**CEST**		
	30 bar	***40 bar***	***63 bar***	***82 bar***	***30 bar***	***40 bar***	***63 bar***	***82 bar***
50	24.39	28.80	32.70	34.77	27.83	31.48	35.38	36.26
48	27.56	31.10	35.21	37.77	30.24	33.77	37.89	40.44
46	29.96	33.70	38.05	40.39	32.64	36.39	40.73	43.06

Table 12 again shows that bagasse moisture content plays a critical role in contributing to CO_2 savings and hence global warming mitigation. It could be seen that an inverse relationship exist between CO_2 savings and bagasse moisture content. As moisture content decreases, savings in CO_2 increases because of the increase in the production of green electricity as already explained in the preceding sections. At the base bagasse moisture content of 50%, the BPST system under the less efficient mill case contributes 24.39 tons/h in CO_2 savings. A 2% drop in moisture leads to about 13% increase in CO_2 savings while at 46% moisture content about 23% increase is recorded. Similarly 8.66% and 17.28% increase in CO_2 savings are recorded under the efficient mill case as bagasse moisture drops to 48% and 46% respectively. It must however be noted that achieving lower bagasse moisture content and hence higher CO_2 savings comes at the expense of extra energy usage in the dewatering and drying mills employed in the sugar milling process.

The effect of turbo-alternator type and operating pressure on CO_2 savings is also shown in Table 12. It is observed that CO_2 savings increase with increasing operating pressure. Changing the turbo-alternator type from a BPST system operating at 30bar pressure to a the advanced CEST system operating at 82bar pressure, increase the CO_2 savings from 24.39 ton/h to 34.77 tons/h at 50% bagasse moisture under the less efficient mill condition. This represents a 42.55% increase in savings. Under the same mill conditions, about 66% increase in CO_2 savings is obtained when bagasse moisture is dropped to 46% and CEST system at 82bar is used instead of the BPST system. Similar trends are observed under the efficient mill condition where at 50% bagasse moisture content, about 30% savings in CO_2 is obtained by changing from the BPST system to 82bar CEST system and about 55% CO_2 savings is obtained at 46% moisture content using 82bar CEST system. Table 12 shows that for the sugar industry to benefit much from carbon credit trading and at the same time meet their energy needs the sugar mills must invest in advanced combustion systems and also strive to reduce the moisture content of the bagasse through improvement in the efficiency of their dewatering systems.

4.4.2 Pure Fast Pyrolysis BMECP Result

Table 13 below gives a summary of the modelling results obtained in this study when using the Pure Fast Pyrolysis BMECP to convert bagasse into energy and energy products. This model was built based on 50% bagasse moisture content only as it is currently the probable moisture content achieved in most sugar mills.

It is seen from Table 13 that the BMECP generates about 124 tons/h and about 153 tons/h of HP steam for the efficient mill case and the less efficient mill case respectively for electricity production and subsequent LP steam extraction to the mills.

Energy for pyrolysis and steam production is provided through the combustion of all the biochar produced in the process hence leaving a net biochar production of zero. However, the pyrolysis process places a limitation on the quantity of heat available for steam production due to its own heat requirement (heat is extracted to the pyrolysis reactor to run the pyrolysis process). Thus to produce enough heat to generate the required amount of steam needed to run the sugar mill operations, a percentage of the total bio-oil (31.2 tons/h) is consumed alongside the biochar. Under the efficient mill condition, 46.5% of the total bio-oil is consumed to meet the steam demand of the mill resulting in final bio-oil product of 16.69 tons/h for storage and future use. About 61% of the total bio-oil is consumed to meet the demands of the less efficient mill leaving final bio-oil product of 12.17 tons/h.

The final electricity outputs from the BMECP after subtraction of parasitic load (electricity consumed in grinding bagasse and for running unit equipment such as pumps and compressors in the pyrolysis process) are 11.05MW and 15.29MW for the efficient mill and the less efficient mill respectively. Although these values are lower than what was generated using combustion, the bio-oil produced is a high value energy product and can be combusted to generate additional electricity. The difference seen in the two values is because of the higher flowrate of steam going to the turbine of the less efficient mill. This is however compensated for by the bio-oil since the efficient mill produces more final bio-oil product than the less efficient mill; 16.69 tons/h against 12.19 tons/h.

Table 13: Summary of Pure Fast Pyrolysis BMECP Results

	Sugar Mill Type	
	Efficient Mill	*Less Efficient Mill*
Turbo type	40bar CEST	40bar CEST
HP steam, tons/h	123.73	152.90
tons/tch	0.41	0.51
tons/ton bagasse	1.53	1.89
Mill Energy Consumption		
Steam @ 2bar, 120 C	120.00	150.00
Mill electricity, MW	6.00	6.00
kW/tch	20.00	20.00
Electricity Generation		
Total output from BMCEP, MW	11.05	15.29
kW/tch	36.85	50.96
Export electricity, kW/tch	16.85	30.96
Pyrolysis Products		
Bio-oil, tons/h	16.69 (31.20)[a]	12.17 (31.20)[a]
Net biochar, tons/h	0.00 (7.73)[b]	0.00 (7.73)[b]
Energy efficiency		
Electrical, %	6.62	9.16
Overall, %	87.65	85.86
CO_2 savings		
tons/h	41.30	38.35
Total biomass, tons/h	81.00	81.00

[a] Value in bracket indicates the total bio-oil produced.
[b] Value in bracket indicates total biochar produced from the pyrolysis reactor.

The low electricity outputs result in lower electrical efficiencies as seen in Table 13 however, the overall system efficiencies are high. These high overall system efficiencies are as a result of the contribution from the final bio-oil product energy to the total energy output of the process and also the inclusion of the energy content of the steam going to the mills. When converted to electrical power assuming a 10% loss in input energy during combustion and 40% conversion of steam energy to mechanical energy in a steam turbine (www.mpoweruk.com), the final bio-oil product could generate an additional 30.2MW and 22MW of electricity for the efficient mill and the less efficient mill respectively. This would increase the respective electrical efficiencies from 6.62% and 9.16 % to

24.72 and 22.36%. However, the overall system efficiencies would be reduced because of the loss in conversion efficiency encountered when converting thermal energy to electrical energy.

Based on the final bio-oil products and the electricity generated, the CO_2 savings as estimated for the efficient mill and the less efficient mill are 41.3 tons/h and 38.35 tons/h respectively. It is shown that improvement in mill efficiency result in about 8% increase in CO_2 savings when using the Pure Fast Pyrolysis BMECP.

4.4.2.1 Effect of Pyrolysis Product Usage for Heat Generation on Pure Fast Pyrolysis Efficiency

Considering only final pyrolysis products, the performance of only the pyrolysis section of the Pure Fast Pyrolysis BMECP modelled in this study was compared to the result from a previous study by Leibbrandt (2010) in Table 14. Leibbrandt (2010) modelled a Pure Fast Pyrolysis process solely for the conversion of bagasse into bio-oil and biochar. The comparison depicts the effect the use of pyrolysis products to supply the energy demand of the sugar mill has on the efficiency of the pyrolysis process. This efficiency otherwise referred to as Liquid plus Thermal Products efficiency by Leibbrandt (2010) was estimated according to the following modified equation 6:

$$\eta_{thermal} = \frac{\sum \mathrm{E}_{th} \text{ final pyrolysis products}}{\mathrm{E}_{th} \text{ Biomass}} \qquad 6$$

Where, E_{th} = thermal energy (MW) in total biomass feed input and final pyrolysis products.

Table 14: Effect of Pyrolysis Products Use on Pure Fast Pyrolysis BMECP Efficiency

	This Study		**Leibbrandt (2010)**
	Efficient Mill	***Less Efficient Mill***	
Biomass input, tons/h	81.00	81.00	41.58
Pyrolysis products, tons/h			
Total bio-oil	31.20	31.20	16.24
Final bio-oil	16.70	12.17	16.24
Total biochar	7.73	7.73	3.49
Final biochar	0	0	1.49
Efficiency (based on final products)	50.29	36.67	69.70

In the model of Leibbrandt (2010), all the bio-oil produced was stored as final product while about 57% of the total biochar was combusted to run the process which had no external heat demand as was the case in this study, thus the relative percentage of total final product was high. Table 14 shows that using a part of the pyrolysis products to help meet the energy demand of the sugar mill causes the efficiency of the Pure Fast Pyrolysis BMECP model based on final pyrolysis products to decrease to 50.29% and 36.67% in the case of the efficient mill and less efficient mill respectively instead of the about 70% obtained by Leibbrandt (2010). The analysis shows that sugar mill efficiency is very critical as it has a direct impact on the quantity of final pyrolysis product and hence the efficiency of pyrolysis. Energy integration of the sugar milling process is thus needed if a Pure Fast Pyrolysis BMECP is to be implemented for sugar mill energy production. This would lead to improvement in efficiency and increase the availability of final pyrolysis product, which then could be offered for sale for additional revenue generation for the sugar industry.

4.4.3 Partial Fast Pyrolysis BMECP Result

A summary of the result for this BMECP model is shown in Table 15 below. As explained in chapter three, this BMECP model combusts part of the incoming bagasse feed input sufficient enough to provide both the sugar mill energy demand and the energy for pyrolysis. The remaining bagasse feed is then taken through fast pyrolysis to produce pyrolysis products for sale or future energy generation.

To fully satisfy the energy requirement of the sugar mill, 70% of the input bagasse feed to the BMECP is combusted under the efficient mill scenario while 80% is combusted under the less efficient mill scenario. It is seen from Table 15 that steam generation in both sugar mill cases are just above the requirement of the mill just as was observed under the Pure Fast Pyrolysis BMECP model due to the need to save sufficient bagasse for pyrolysis product production as well the extraction of heat from the combustor for pyrolysis operation.

It is seen from Table 15 that electricity generation from this BMECP after parasitic load subtraction is 16.55 MW for the efficient mill and 20.78 MW for the less efficient mill with corresponding electrical efficiencies of 9.94% and 12.45% and 35.28kW/tch and 49.21kW/tch of exportable electricity respectively. Due to the high steam demand of the less efficient mill, high steam flowrate is also directed to its steam turbine resulting in the 25% increase in its electricity generation from that of the efficient mill. Another possible contribution to this is the small quantity of heat (8.64MW) that is extracted to the pyrolysis reactor of the less efficient as compared the 12.97MW supplied to the reactor of the efficient mill thus making more heat available for steam generation which in turn increases electricity production.

Table 15: Summary of Partial Fast Pyrolysis BMECP Results

	Sugar Mill Type	
	Efficient Mill	***Less efficient mill***
Turbo type	40bar CEST	40bar CEST
HP steam, tons/h	128.72	152.92
tons/tch	0.43	0.51
tons/ton bagasse	1.59	1.89
Mill energy consumption		
Steam @ 2bar, 120 C	120.00	150.00
Mill electricity, MW	6.00	6.00
kW/tch	20.00	20.00
Electricity Generation		
Total output from BMCEP, MW	16.58	20.78
kW/tch	55.28	69.27
Export electricity, kW/tch	35.28	49.27
Pyrolysis Products		
Bio-oil, tons/h	9.36	6.24
Net biochar, tons/h	2.32	1.55
Energy efficiency		
Electrical, %	9.94	12.45
Overall, %	85.09	84.36
CO_2 savings		
tons/h	40.44	38.05
Total biomass, tons/h	81.00 (56.70)[a]	81.00 (64.8)[a]

[a] Value in bracket indicates the quantity of bagasse bypass to the combustor

The hourly bio-oil and biochar production using this BMECP scenario are 9.36 and 2.32 tons respectively for the efficient mill. The less efficient mill produces 6.4 and 1.55 tons of bio-oil and biochar respectively, which are less than the efficient mill values due to the small quantity of bagasse that is pyrolyzed (the high steam demand leaves less surplus bagasse for pyrolysis). This shows that close to 50% increase in both bio-oil and biochar production can be achieved when mill steam usage and hence efficiency is improved from 50% tons/tch (less efficient) to 40% ton/tch (efficient).

Table 15 also shows that high overall system efficiencies are achievable when the Partial Fast Pyrolysis BMECP is used for sugar mill biomass processing into energy and energy products. Under the efficient mill condition, this BMECP technology achieved overall system efficiency of 85%. This system efficiency is 1% higher than that obtained under the less efficient mill condition and is expected to increase further as mill steam usage is drop further down. This in turn will lead to an increase in the production of pyrolysis products for sale while also improving the overall efficiency of bagasse conversion.

4.4.4 Comparison of BMECP Technologies

The performances of the three BMECP technologies studied in this work are compared below in Table 16 and Table 17 for the less efficient and efficient mill cases respectively. The comparisons were done based on the same bagasse feed input rate and 50% moisture content (current moisture content commonly found at most sugar mills). It can be seen from both Table 16 and Table 17 that while the two pyrolysis-based BMECP technologies could only produce steam that is just about a little above the quantity required by the sugar mills, their combustion counterparts could produce steam quantities very much over and above the requirement of the mill. This is especially the case when considering the efficient mill which needs only 120 tons/h of LP steam for it operation. At the same HP steam pressure condition, about 160 tons/h of steam can be produced when using Combustion in both cases of the sugar mill, while the Pure Fast Pyrolysis BMECP and the Partial Fast Pyrolysis BMECP can only produce about 153 tons/h under the less efficient mill condition and about 129 tons/h and 124 ton/h respectively under the efficient mill condition. The cause of this limitation to steam generation is the extraction of some of the thermal heat energy from the respective combustors of the pyrolysis-based BMECPs to supply the energy need of their pyrolysis reactors.

Table 16: Comparison of BMECP Technologies Performances for Less Efficient Mill

	Combustion				**Partial Fast Pyrolysis**	**Pure Fast Pyrolysis**
	BPST	CEST			CEST	CEST
	30 bar	40 bar	63 bar	82 bar	40 bar	40 bar
Steam, tons/h	161.73	159.56	153.15	150.59	152.92	152.90
Electricity						
Total, MW	21.40	25.26	28.68	30.50	20.78	15.29
Export, kW/tch	51.32	64.20	75.60	81.68	49.27	30.96
CO_2 savings, tons/h	24.39	28.80	32.70	34.77	38.05	38.35
By-product, tons/h						
Bio-oil	-	-	-	-	6.24	12.17
Biochar	-	-	-	-	1.55	0.00

The trend in steam generation is also seen for both total electricity production and exportable electricity. From Table 17, compared to a total electricity of 27.61 MW and surplus electricity of 72 kW/tch produced by the Combustion technology at 40 bar pressure, the Partial Fast Pyrolysis technology produces only 16.58 MW total electricity and 35.28 kW/tch surplus electricity while the Pure Fast Pyrolysis technology also produces 11MW total electricity and 16.85 kW/tch exportable electricity at the same condition. The use of advanced Combustion technology result in even higher electricity production; 30.50 MW (see Table 16) and 31.81 MW (see Table 17) of total electricity produced under the less efficient mill and the efficient mill conditions respectively. This trend in electricity production is partly due to the limitation placed on steam generation by the pyrolysis process and also the consumption of part of electricity produced by auxiliary equipments in the pyrolysis processes (electricity is consumed to run equipment such as pumps and compressors and also to grind bagasse to <2mm particle size).

Table 17: Comparison of BMECP Technologies Performances for Efficient Mill

	Combustion				**partial Fast Pyrolysis**	**Pure Fast Pyrolysis**
	BPST	CEST			CEST	CEST
	30 bar	40 bar	63 bar	82 bar	40 bar	40 bar
Steam, tons/h	161.73	159.56	153.15	150.59	128.72	123.73
Electricity						
Total, MW	24.42	27.61	31.03	31.81	16.58	11.05
Export, kW/tch	61.40	72.04	83.44	86.02	35.28	16.85
CO_2 savings, tons/h	27.83	31.48	35.38	36.26	40.44	41.30
By-product, tons/h						
Bio-oil	-	-	-	-	9.36	16.69
Biochar	-	-	-	-	2.32	0.00

By comparing the Pure Fast Pyrolysis process to the Partial Fast Pyrolysis process, it is seen from both Table 16 and Table 17 that the electricity production from the Partial Fast Pyrolysis BMECP is always higher than that of Pure Fast Pyrolysis BMECP. This is because more electricity is spent to grind bagasse in the Pure Fast Pyrolysis BMECP where all bagasse input undergoes pyrolysis than is spent in the Partial Fast Pyrolysis BMECP where only a fraction of the bagasse input is taken through pyrolysis.

Despite the low electricity production from the pyrolysis-based BMECPs, these technologies produce, in addition to electricity, high value bio-oil and biochar, which can further be processed to electricity immediately or during offseason for sale to bring in added income.

It is also shown from Table 16 and Table 17 that the two pyrolysis-based process technologies contribute more towards the mitigation of global warming and hence are more environmental friendly than combustion technology. This is so due to the storage of carbon in pyrolysis products.

From the analysis above the combustion technology proves to the most appropriate option when on-season or immediate electricity production from sugar mill biomass is required as it produces more electricity than the two pyrolysis-based technologies, especially when advanced steam/electricity generating turbo-alternator operating at higher pressure is used. However this technology contributes less towards CO_2 savings and produces no additional high energy valued products for off-season power production.

CHAPTER FIVE

Economic Analysis

5.1 Introduction

Among the objectives of this study is the development of economic models for the BMECP technologies simulated in Aspen Plus® which will allow for the comparison of the different process models on the basis of economic viability, since economic viability forms the baseline of every manufacturing enterprise.

For this study, the economic model analyses for the various BMECP plants were done using the Aspen Process Economic Analyser V7.1 software developed by Aspen Technology, Inc. In all the economic models the Total Capital Investment (TCI), total operating cost and economic viability of each BMECP technology were determined based on an nth plant approach. This approach assumes that the technologies applied in the various BMECP models have attained application in already established commercial plants and are well understood, even though fast pyrolysis application for steam and electricity production has yet to attain commercial status. This implies cost associated with a pioneer plant such as those for risk financing and longer start-ups are not included in the analysis. The TCI was estimated first based on total installed cost of equipment sourced from vender quotations and literature as well as from the equipment cost library of the software. The operating cost was estimated next and the two together were used in a discounted cash flow analysis to establish profitability. This section describes in detail the procedure and the assumptions used in developing the economic models after which results are presented. The effects of changes in major economic input parameters were also studied through sensitivity analysis.

5.2 Total Capital Investment (TCI)

The total capital cost associated with each process model was determined using the Aspen Process Economic Analyzer V7.1 software. Aspen Plus® process model flowsheets together with their mass and energy balances were imported into the software. Standard equipment were then mapped to standard equipment in the equipment model library of Aspen Process Economic Analyzer and sized using the sizing expert tool of the software as well as standard sizing methods outlined in the Chemical Engineer’s Handbook (Perry and Green, 2008).

The base cost year used for the analysis was 2008 (i.e. the base cost year which comes with the V7.1 version of the analyzer used in this study) hence the price of all equipment were indexed to 2008 dollars and then adjusted to the start date of basic engineering to obtain the TCI in the project year using Equation 7 below (Aspen Technology, Inc., 2009):

$$C_{at} = C_t \times \left[1 + \left(t_{diff} \times e\right) / \left(w \times 7 \times 100\right)\right] \tag{7}$$

Where C_{at} = adjusted total capital cost, C_t = total capital cost, t_{diff} = time difference between system cost base date and start date for basic engineering, e = project capital escalation (user defined) and w = number of weeks per period.

Standard equipment such as pumps, heat exchangers, compressors, etc. were costed by the software based on user defined specifications. Table 18 below shows the general specifications used for the economic models in Aspen Process Economic Analyzer.

Table 18: General specifications used for economic models

Item	**Specification**
Process description	Redesigned process
Process complexity	Typical
Process control	Digital
Project Information	
Project location	Africa
Project type	Grass roots/clear field
Contingency	
Estimated start date of basic engineering	1st July, 2012
Soil condition around site	Gravel
Equipment Specification	
Pressure vessel design code	ASME
Vessel diameter specification	ID
P and I design level	FULL

The contingency expressed as a percentage of TCI is generated by a combined contribution from the project type, process complexity and the process description. The process description also sets the design allowance for equipment of which the material costs are generated by the software (Aspen Technology, Inc., 2009). The project location affects the cost associated with domestic and ocean freight (% of material cost), taxes/duty (% of material cost) and equipment rotating spares percentage. The power distribution system of the plant is also defined by the project type while the site soil condition affects the type of pile foundation required for equipment (Aspen Technology, Inc., 2009).

For non-standard equipment such as the biomass combustor/boiler system, pyrolysis equipment and feed preparation equipment which come in unit blocks, Aspen Process Economic Analyzer cannot determine their prices accurately. Costs for such equipment together with their installation factors were determined from published cost data used in other techno-economic analysis studies found in literature and from vender quotes and added to the quoted equipment cost library of the software for total capital cost estimation.

Since quoted equipment costs are quoted for different size capacities other than those used in this study, Equation 8 below was used to resize and to obtained the new cost corresponding to the capacities used in this study where, n is the exponential factor in the range of 0.6 – 0.8 (Peters and Timmerhaus, 2003).

$$\text{New Cost} = \text{Original Cost}\left(\frac{\text{New Capacity}}{\text{Old Capacity}}\right)^{n} \qquad (8)$$

Also since quoted equipment costs are given in different base cost years, the cost for such equipment were adjusted to the 2008 base cost year used in this analysis using the Chemical Engineering Plant Cost Index (CEPCI) according to Equation 9 below (Peters and Timmerhaus, 2003):

$$\text{New Cost} = \text{Base Cost}\left(\frac{\text{CEPCI in New Year}}{\text{CEPCI in Base Year}}\right) \qquad (9)$$

Table 19 lists the CEPCI data used in this study and are also plotted in Figure 34. Data is sourced from Humbird et al. (2011) and the values from the year 2011 to 2015 are obtained through linear regression analysis.

Table 19: Chemical Engineering Cost Indices used in Calculation

Year	Index	Year	Index
1990	357.6	2003	402.0
1991	361.3	2004	444.2
1992	358.2	2005	468.2
1993	359.2	2006	499.6
1994	368.1	2007	525.4
1995	381.1	2008	575.4
1996	381.7	2009	521.9
1997	386.5	2010	550.8
1998	389.5	2011	481.7
1999	390.6	2012	483.9
2000	394.1	2013	486.1
2001	394.3	2014	488.4
2002	395.6	2015	490.6

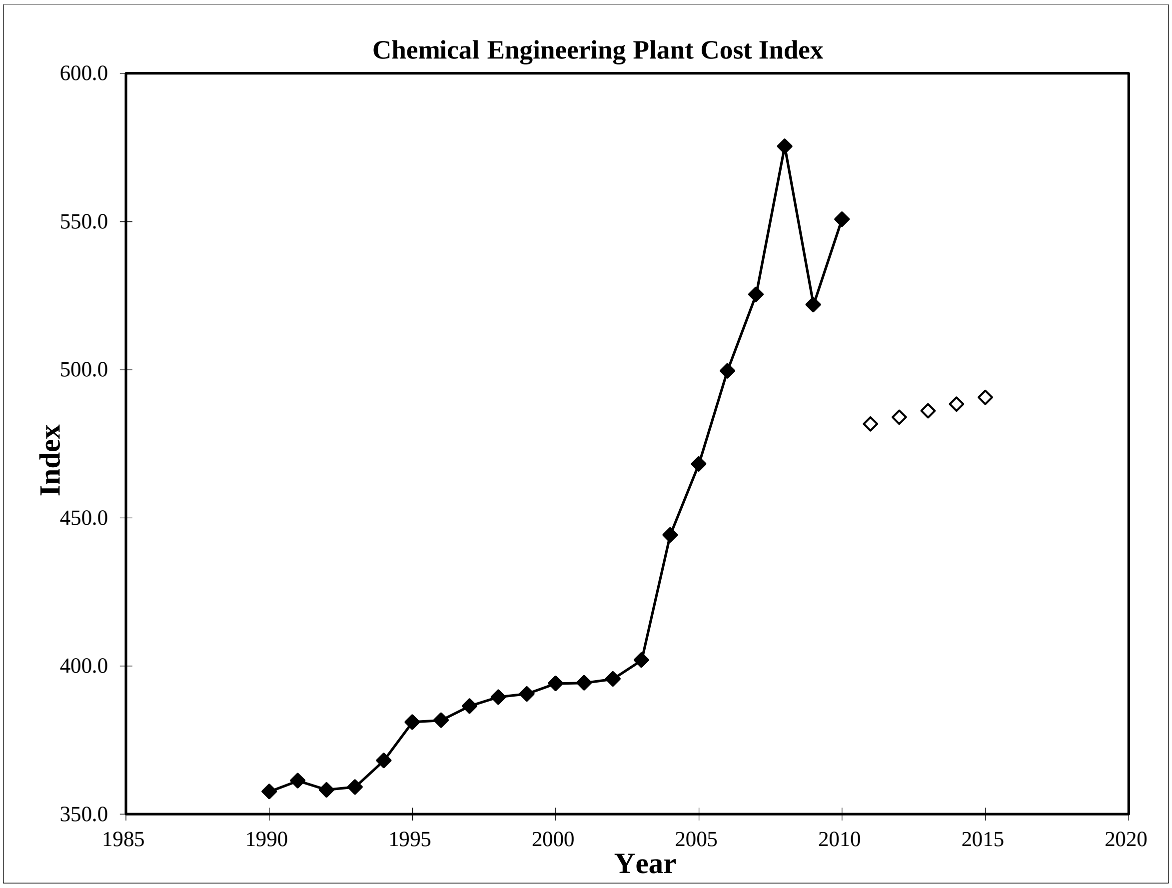

Figure 34: Chemical Engineering Plant Cost Indices (Redrawn from Humbird et al., 2011)

5.3 Operating Cost

Operating cost is classified into two: fixed operating cost and variable operating cost.

5.3.1 Variable Operating Cost

Variable operating cost are those cost items whose value are directly dependent on the operating capacity/operating rate of the plant at any given time period (Peters and Timmerhaus, 2003). They include cost for raw materials, charges relating to waste handling and treatment, by-product credits, utilities and chemicals and credits from product sales.

For this study, a fixed bagasse feed rate of 81 ton/h (50% moisture) was assumed for all the economic models. Products and by-products rates and utility usage were determined through mass and energy balances developed in Aspen Plus®. Table 20 lists the cost associated with the raw material, products and by-products streams as used in this study for the economic evaluation.

Table 20: Cost Data for Feedstock, Products and By-products

Stream	Cost
Bagasse	56 $/dry ton
Electricity	0.248 $/kWh
Biochar	120 $/ton
Bio-oil	170 $/ton

The bagasse price was estimated using the formula developed by Jenkins (1997) and assuming a 50 km radius travelling distance to and from the mill for delivery (Botha and Blottnitz, 2006). The formula taken from the work of Pippo et al. (2009) assumes that bagasse cost can be estimated through the cost of fuel oil since it is used as a substitute for #6 fuel oil in boilers. However, in the context of South Africa, mills usually use coal as an alternative fuel in place of bagasse. Hence the formula was modified to accommodate the properties (LHV and price) of coal instead of those of #6 fuel oil and the cost of bagasse determined as the value of coal of equivalent energy content. The price of $56/dry ton (28/wet ton) obtained compares very well with values quoted in literature for biomass cost. Pippo et al. (2009) obtained a bagasse price of $27.7/wet ton (50% moisture) in their study. Similarly Humbird et al. (2011) estimated a feedstock price of $58.5/dry ton for corn stover, which is an agricultural residue with similar properties as sugar cane bagasse. In energy terms, the $56/dry ton price translates to approximately $7.47/GJ which is also similar to the $7.7/GJ used by the National Energy Regulator of South Africa (NERSA, 2011)) to estimate the Levelised Cost of Electricity (LCOE) for cogenerated electricity using sugar cane bagasse as fuel.

The electricity cost was assumed to be the same as the LCOE estimated by NERSA for bagasse derived electricity in consultation with the SMRI. It was assumed that biochar will replace coal as boiler fuel. By comparing the energy value of char 24.65 MJ/kg (Hugo, 2010) to the energy value of export grade bituminous coal 24.7-26 MJ/kg (Eberhard, 2011), it is seen that biochar is equivalent to coal on energy basis hence its price was set to be the same as that of coal which, sold for an average price of $120/ton between July 2011-December, 2011 (www.indexmundi.com). A similar assumption was made by Leibbrandt (2010).

Since the bio-oil in this study will be used as combustion fuel, it was also assumed that it will replace #6 fuel oil, a commonly used industrial heating fuel. However, due to the lack of a readily established market for bio-oil, its price was assumed to be 70% of that of #6 fuel oil. Fuel oil #6 sold for $2.159/gal ($13.375/GJ) in 2011 (Energy information Administration, 2012). Given the soaring increases in crude oil price over the past years, it is not envisioned that the price of #6 fuel oil (a derivative from crude oil) will drop significantly below the 2011 price level, thus the price of

$2.159/gal ($13.375/GJ) was used which translates into a bio-oil price of $170/ton and used in this study for the economic modelling.

5.3.2 Fixed Operating Cost

Fixed operating costs are not related directly to the production rate of the plant and therefore are incurred fully irrespective of the production rate of the plant (Peters and Timmerhaus, 2003). They include cost for labour, maintenance, operating charges, general and administration (G and A) expenses and other plant overhead items. In all the economic models, labour costs were estimated by Aspen Process Economic Analyzer. The required workforce to operate and supervise the plant per shift (hours) was determined and then adjusted to account for the total number of hours the plant operates for each period and then multiplied with the respective unit labour cost to obtained the total labour cost. System default values/percentages (see Table 21) consistent with those specified by Peters and Timmerhaus (2003) were accepted for operating charges, overheads and G and A expenses. The operating charges expressed as percentage of total operating labour include cost for operating supplies and laboratory. Plant overheads (% labour and maintain cost) include cost for services, payroll overhead, facilities, etc incurred during production. G and A expenses consist of general and administrative cost associated with R&D, product distribution, sales cost and salaries/expenses incurred during production (Peters and Timmerhuas, 2003). G and A expense is expressed as a percentage of subtotal operating cost (Aspen Technology, Inc., 2009).

Table 21: Operating Cost Input Parameters used in Aspen Process Economic Analyser

General	
Plant Overhead	50 (%/yr)
Operating Charges	25 (%/yr)
G and A Expenses	8 (%/yr)
Unit Labour Cost	
Operator	20 ($/person/hour)
Supervisor	35 ($/person/hour)

5.4 Profitability Analysis

Knowledge of the economic viability/feasibility of a project is a prerequisite tool investors look out for before investing money into any project. Economic viability of the various BMECP technologies modelled in this study were therefore evaluated through a discounted cash flow rate of return (DCFROR) analysis after the determination of TCI and total operating cost. In all cases, the following indictors: net present value (NPV), internal rate of return (IRR), payout period (PO) and profitability index (PI) were used to establish the profitability of the models. These indicators are briefly discussed below based on the definitions given by Aspen Technology, Inc., (2009) unless otherwise stated:

- The NPV is an indication of how much all net earnings obtained through a period n is worth currently. It is obtained through a discounting technique where the net earnings in current time period n is added to those from previous periods and discounted back to the first time period at a given discount rate to give the NPV for that period n. NPV accounts for the time value of money and its sign at any given period of analysis determines the profitability of a project. A positive NPV indicates a project is profitable while a negative value indicates the opposite. NPV of zero implies a project neither makes gains or losses and is referred to as the break-even point.
- IRR is the discount rate at which NPV is zero. It is that interest rate at which money can be borrowed and invested into a project and still break-even at the end of a projects life after tax deductions. IRR greater than prevailing interest/lending rate would yield a positive NPV and result in a profitable project whereas a value less than prevailing rate would yield a negative NPV (non-profitable project). Thus the more an estimated IRR is greater than the current interest rate the more the return on investment.
- PO is the time period needed to recover and repay the initial original capital outlay in a project. The PO unlike the NPV and IRR does not account for the time value of money and does not involve a discounting technique (Gnansounou et al., 2005).
- PI is the ratio of the present value of cumulative cash inflows to the present value of cumulative cash outflows for a period n. It is the relative relationship between the present value of benefits and the present value of cost. PI>1 is an indication that a project is profitable while PI<1 indicates otherwise.

DCFROR analysis requires the specification of certain investment parameters which include the plant economic life, tax rate, interest/discount rate, plant operating hours or plant attainment, the depreciation method and the duration of the start-up period among others. The specification for these parameters is shown in Table 22. As shown in Table 22, economic model analysis for all the BMECP technologies were done over a plant economic life period of 20 years using the straight line depreciation method which allows for evenly depreciation of a project over its economic life (Aspen Technology, Inc., 2009). The straight line depreciation method was chosen because is the most commonly used method (Peters and Timmerhaus, 2003; Aspen Technology, Inc., 2009). The 20 year economic life period was used to commensurate with what is commonly used in literature (Kazi et al., 2010; Ringer et al., 2006; Gnansounou et al., 2005; Rhodes and Keith, 2005; Swanson et al., 2010; Anex et al., 2010) for technoeconomic analysis of various biomass to energy projects.

Plant start-up time was set at 25% of the respective construction time for each plant in accordance with the suggestion of Perry and Green (2008) while the tax rate used in all the economic models was set at the highest tax rate at which companies operating in South Africa are taxed (Page Accounting, 2011; GranThorton South Africa, 2011; Price Water Coopers, 2009; South African Revenue Service, 2009). The salvage value (% of initial capital investment) falls within the 5-15% range recommended by Peter and Timmerhaus (2003).

Table 22: Investment Analysis Parameters used for Economic Modelling

General Investment Parameters	
Economic Life of Project	20 years
Depreciation Method	Straight Line
Tax Rate	28 %/year
Interest Rate/Desired Rate of Return	10 %/year
Salvage Value	10 % of Initial Capital Cost
Escalation Parameters (%/year)	
Project Capital	8
Products	8
Raw Materials	8
Operating and Maintenance Labour	8
Utilities	8
Project Capital Parameters (%/year)	
Working Capital	5
Facility Operating Parameters	
Operating Mode	Continuous Processing – 24 h
Operating Hours per Year	8000
Length of Start-up Period	25 % of Construction Time

For interest rate/discount rate, Short et al. (2005) are of the view that “In the absence of statistical data on discount rates used by industrial, transportation and commercial investors for investments with risks similar to those of conservation and renewable energy investments, it is recommended that an after tax discount rate of 10%...be used.” Based on this recommendation, a 10% interest rate/discount rate was used in all the economic models. A similar percentage was used in other biomass to energy technoeconomic studies (Humbird et al., 2011; Anex et al., 2010; Swanson et al., 2010; Kazi et al., 2010; Rhode and Keith, 2005). This assumption is valid in the context of South Africa given the fact that the BMECP plants modelled in this study if found to be economically viable will be sited in South Africa where the prime interest rate (PIR) has remained constant at 9% since November, 2010, as shown in Figure 35 below (South African Reserve Bank, 2012). The PIR is the benchmark rate used by commercial banks to lend out loan to the public. Also the 10% assumed rate used is only slightly higher than the 9.8% used by NERSA (2011) to estimate the LCOE of bagasse derived electricity in South Africa.

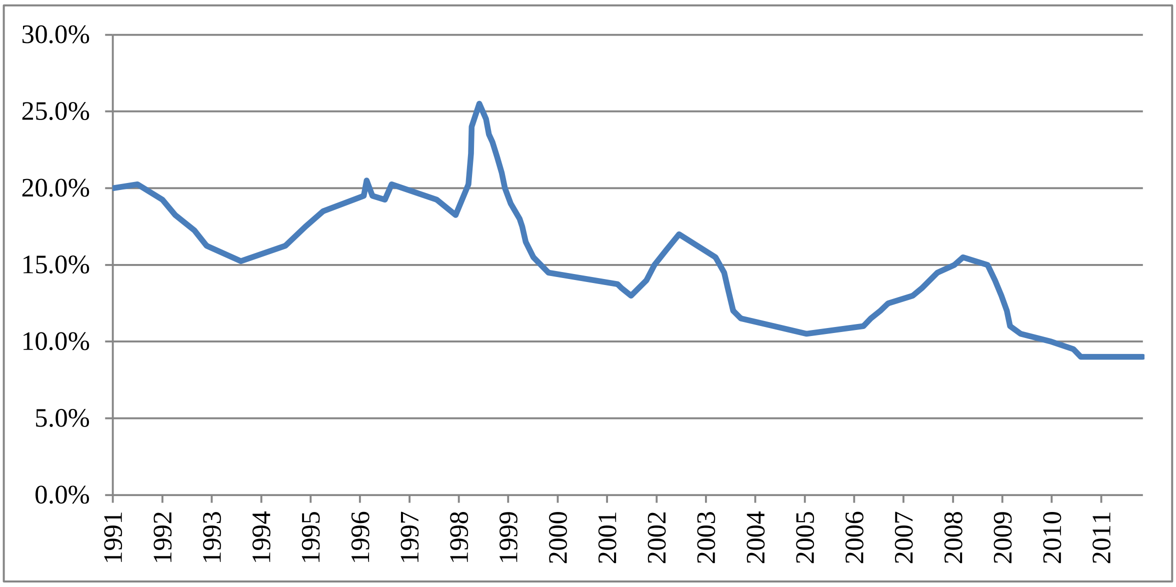

Figure 35: Historical trend of South African interest rates (South African Reserve Bank, 2012)

The escalation parameters are the percentage rates at which project capital expenses, products sales revenue, cost of raw material, operating and maintenance labour cost and utilities costs are increased annually (Aspen Technology, Inc., 2009). The respective values used in this study are typical values applied by the South African sugar industry (Leibbrandt, 2010). Working capital was estimated at 5% of TCI and is in the range set by Garrett (1989). It is the capital required to operate a facility until there is sufficient revenue from product sales to cover cost (Aspen Technology, Inc., 2009).

5.5 Economic Modelling Results and Discussion

Summary results of the economic modelling of the various BMECP plants studied in this project is presented in Table D1.1 and Table D1.2 in Appendix C1 for both the efficient and the less efficient mill conditions respectively, with all cost data reported in $US (year 2008). For each BMECP economic results, information presented in the above mentioned Tables include investment cost (Total Capital Investment), operating costs, products sales revenues as well as summary of the yield performance of the plant and are discussed below.

To allow for effective comparison of the various BMECP technologies on a common basis, economic results of the combustion based BMECPs are given for only the 50% moisture bagasse feed input; the same moisture content used for the two pyrolysis based models. It must however be noted that the complete economic results for all the BMECP models simulated in Aspen Process Economic Analyzer together with the Excel spreadsheet used for the cost estimation are contained on a CD attached to this thesis. Also contained on the CD are detailed investment analysis reports and decision analyzer workbooks (contains cash flow statements and project schedule plans useful for sensitivity analysis) for all the BMECP plants studied in this work.

5.5.1 Investment Costs

Figure 36 and Figure 37 below show the breakdown of the TCI requirement of the various BMECP plants under the efficient and less efficient mills conditions respectively (see also Appendix D2 for tables). A comparison of TCI shows that biomass combustion using 30bar BPST system results in the lowest TCI, costing $68.5 million under the efficient mill (see Figure 36) and $67.5 million under the less efficient mill (see Figure 37). TCI costs for the combustion CEST systems are seen to be significantly greater than the BPST system. The cost increases as the operating pressure of the CEST system is increased from 40bar to 82bar. For example, while there is about 30.75% increment in capital cost for the 40 bar CEST, the capital cost more than double for a 82bar CEST system (103.77% increment) under the efficient mill condition as shown in Figure 36. Similar observations can also be seen under the condition of the less efficient mill. This observation is due mainly to the high cost associated with condensing extraction systems as well as high metallurgy cost associated with high pressure and temperature (directly related to pressure).

Capital investment cost breakdown shows that for the combustion based BMECP plants under both mill conditions, the biomass combustor and steam boiler accounts the most towards the overall TCI (see Figures 36 and 37), with percentage contributions of 82, 78, 82.6 and 85.3 for the 30bar BPST, 40bar CEST, 63bar CEST and 82bar CEST systems respectively under the efficient mill and 83.4, 79.2, 83.3 and 85.7 respectively for the 30bar BPST, 40bar CEST, 63bar CEST and 82bar CEST systems under the less efficient mill condition.

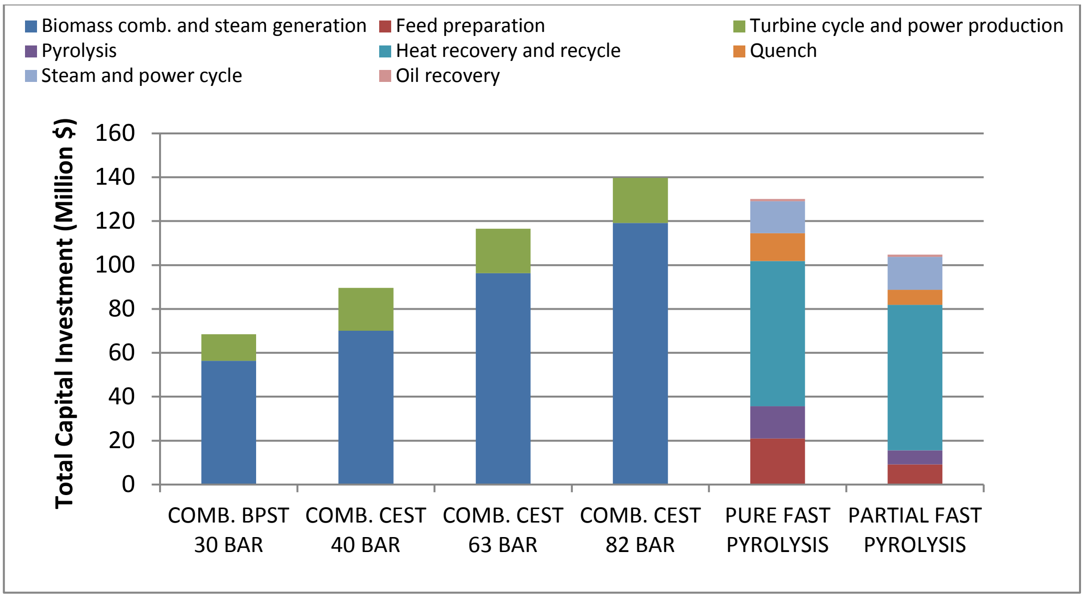

Figure 36: Breakdown of Total Capital Investment for BMECP Models Under Efficient Mill Condition

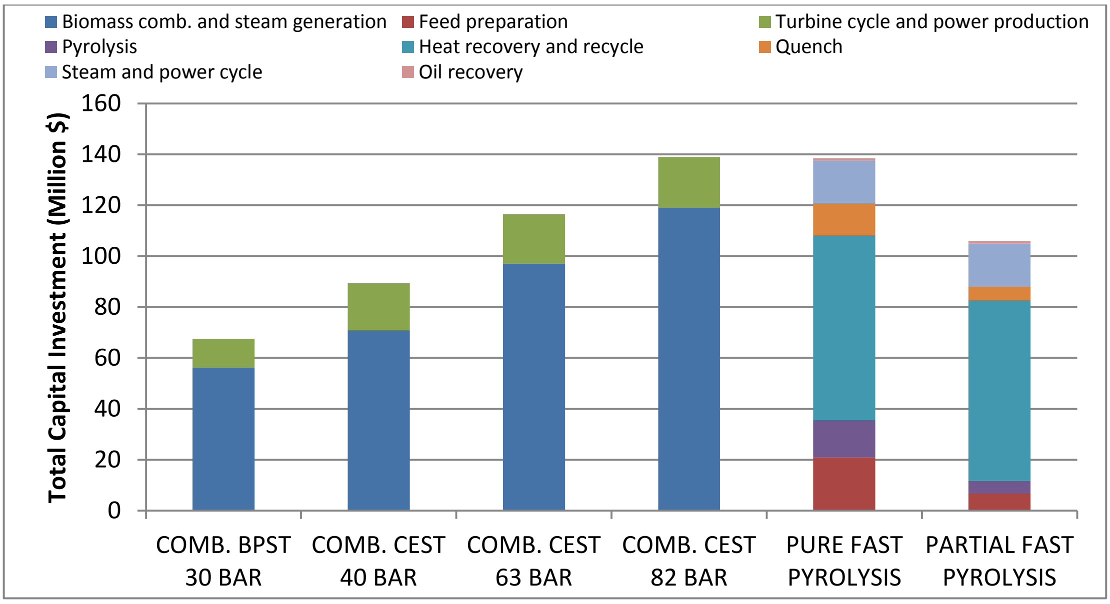

Figure 37: Breakdown of Total Capital Investment for BMECP Models Under Less Efficient Mill Condition

The TCI required for the Partial and Pure Fast Pyrolysis models are $104.72 million and $130.10 million respectively under the efficient mill and $105.94 million and $138.53 million respectively under the less efficient mill. These costs are greater than those of the combustion based BMECPs except for the two advanced CEST systems operating at 63bar and 82bar pressures. The additional unit operation equipment required by the two pyrolysis based processes for bio-oil and biochar production aside those needed for steam and electricity generation (as used in the combustion processes) account for this difference. For this study, a 40bar CEST system was used for the steam and electricity generation section of the pyrolysis based BMECP process modelling which, is a possible reason why their overall total capital investment cost are lower than those of the 63bar and 82bar advanced combustion systems. The installation of advanced steam and electricity generation systems would therefore make the TCI of both the Partial and Pure Fast Pyrolysis BMECP plants to also exceed those of the two advanced combustion based BMECP plants.

The least contributor to the TCI of the pyrolysis based BMECP plants is the oil recovery section contributing 0.81 and 0.93% towards the TCI of the Partial and Pure Fast Pyrolysis plants respectively under the efficient mill and 0.70 and 0.91% towards the TCI of the Partial and Pure Fast Pyrolysis under the less efficient mill. As observed for the combustion based BMECP plants, the heat recovery section (containing the combustor and the steam boiler) again contributes the largest percentage towards the overall TCI of both pyrolysis based BMECP models. However, the contribution towards the TCI of the Partial Fast Pyrolysis BMECP is greater than that of Pure Fast Pyrolysis BMECP; 63.4% versus 50.9% under the efficient mill and 67.0% versus 52.4 % under the less efficient mill. The difference is due to the fact that in the Partial Fast Pyrolysis plant, raw sugar cane bagasse is combusted together with recycled pyrolysis gases in the combustor while the combustor of the Pure Fast Pyrolysis plant uses fine biochar and/or bio-oil together with recycled pyrolysis gasses as fuel. Since bagasse has low bulk density and hence occupies a larger volume than biochar, a larger sized combustor is required for the Partial Fast Pyrolysis BMECP plant and this leads to an increase in the total installed cost of the combustor due to an increase in material cost.

Because the Pure Fast Pyrolysis BMECP consumes all available bagasse for pyrolysis, the contribution of its feed preparation section is greater than that of the Partial Fast Pyrolysis BMECP plant as shown in both Figure 36 and Figure 37. In Partial Fast Pyrolysis only surplus bagasse (after separation of the quantity required to supply the energy need of the sugar mill to the combustor) is taken through pyrolysis hence cost associated with its feed drying and grinding section (necessary requirement for pyrolysis) are lower than in the Pure Fast Pyrolysis BMECP model as a result of the small quantities of bagasse processed for pyrolysis. The contribution of feed preparation towards the TCI of Partial Fast Pyrolysis is 8.7 and 6.5% under the efficient mill and less efficient mill conditions respectively whereas contributions of 16.1 and 15.1% are made toward the TCI of the Pure Fast Pyrolysis BMECP under the same respective mill conditions. The same reason explains why the cost contributions of other process flowsheet areas such as pyrolysis and quench toward the TCI and hence the overall TCI of the Partial Fast Pyrolysis BMECP plants are lower than those of the Pure Fast Pyrolysis BMECP plants as can be seen in Figure 36 and Figure 37.

The effect of sugar mill efficiency on TCI could also be observed when capital cost values for the various BMECP models in Figure 36 (efficient mill) are compared with those in Figure 37 (less efficient mill. For the combustion based BMECP models, TCI under the efficient mill are 1.51, 0.29, 0.13 and 0.46% greater than the TCI under the less efficient mill for the 30bar BPST, 40bar CEST, 63bar CEST and 82bar CEST systems respectively. These increments are due to the increased electric power produced by these BMECP models under the efficient mill condition (refer section 4.3.4) which in turn increases the cost associated with the steam turbine and generator systems since these equipment are cost/sold based on their electrical output. As observed, the percentage changes in TCI of these combustion based BMECP plants are not very significant because HP steam production are the same for these plants under both sugar mill conditions as modelled in this study, thus the biomass combustor and boiler costs which account for a larger percentage of their total capital investments are the same. The opposite however, is observed for the two pyrolysis based BMECP plants. By comparing the TCI of the Partial Fast Pyrolysis and Pure Fast Pyrolysis models in Figure 36 and Figure 37 it could be seen that the implementation of these two pyrolysis based technologies under the less efficient mill condition comes with a higher TCI that under the efficient mill. For instance, the Pure Fast Pyrolysis BMECP cost about $139 million under the less efficient mill condition which is 7.25% greater than it cost under the efficient mill. This is also due to the high quantity of electricity generated by these pyrolysis based models under the less efficient mill than under the efficient mill (see Tables 16 and 17 – section 4.4.4). As stated in chapter three, these two pyrolysis based models are designed to generate only the required steam needed by the sugar mill with little or no excess. Hence more HP steam is generated for the less efficient mill than the efficient mill which in turn increases the electricity production under the less efficient mill (due to increased flowrate of HP steam to the steam turbine) and subsequently the cost of the power generation system as explained above.

5.5.2 Operating Cost

Total operating cost of the combustion based BMECP models, the Partial Fast Pyrolysis and the Pure Fast Pyrolysis BMECP models are shown in Figure 38 and Figure 39 for the less efficient and the efficient mill conditions respectively. Total operating cost ranges from a low of $21.09 million for biomass combustion using a 30bar BPST to $22.97 million for 82bar CEST combustion system under the less efficient mill condition. The costs under the same mill condition are $23.53 million for the Partial Fast Pyrolysis and $23.31 million for the Pure Fast Pyrolysis model (see Figure 38). Under the efficient mill condition, the ranking order as observed for TCI is again depicted, with the 82bar CEST combustion system having the highest operating cost of $25.43 million while the 30bar BPST system has the lowest operating cost of $21.01million. Operating cost in both Figure 38 and Figure 39 is seen to increase gradually for the combustion based BMECP as operating pressure is changed from 30bar to 82bar. This trend in total operating cost is mainly due to the increase in operating labour and maintenance cost and operating charges that comes with high pressure and high temperature operation. As an example, while operating labour and maintenance cost and operating charges are \$0.77 million and \$0.15 million respectively for the 30bar BPST system under the efficient mill condition, those for the 82bar CEST system are \$3.16 million and \$0.66 million respectively (refer to Table D1.1 of Appendix D1 and Figure 39).

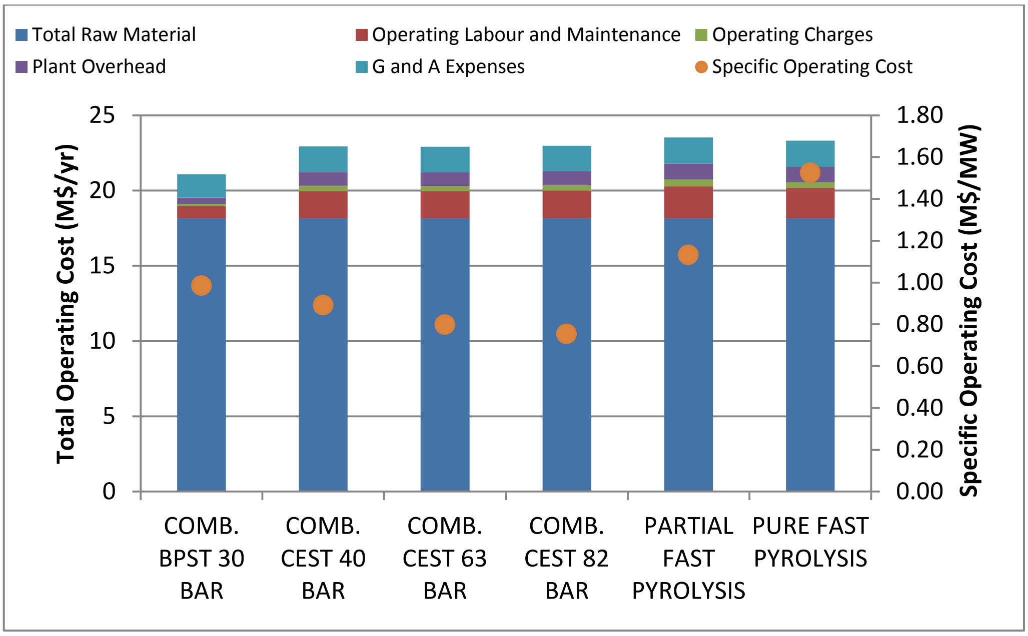

Figure 38: Total and Specific Operating Cost for BMECP Models (Less Efficient Mill)

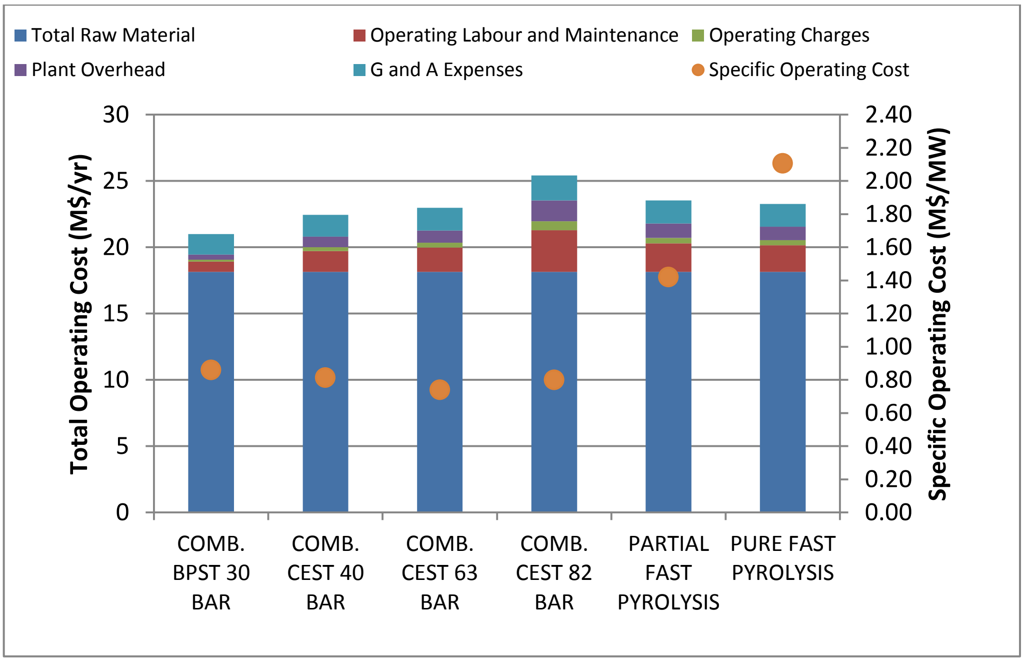

Figure 39: Total and Specific Operating Cost for BMECP Models (Efficient Mill)

Total operating costs were also estimated in this study on a specific basis herein referred to as 'specific operating cost' (i.e. total operating cost divided by total electricity produced, $/MW) and the results are shown in both Figure 38 and Figure 39. As observed in both Figures, the 30bar BPST BMECP have the highest specific operating cost among the combustion based BMECP models even though it recorded the lowest total operating cost. This is because of the low electricity generated by the BPST system when compared to the CEST combustion systems. The trend as seen in both figures also reveal a gradual decrease in specific operating cost with increasing operating pressure of the CEST system due to the gradual increase in electricity generation with increasing operating pressure as discussed in chapter 4.

From both Figures 38 and 39 one can observe that the total operating cost of both the Partial Fast Pyrolysis and the Pure Fast Pyrolysis BMECP models are similar to those of the combustion based models however their specific operating cost are significantly high. Specific operating cost of the Partial Fast Pyrolysis BMECP and the Pure Fast Pyrolysis BMECP models under the less efficient mill condition (Figure 38) are 25.5 and 53.5% respectively higher when compared to that of the 30bar BPST (the highest among the combustion based systems) while their total operating cost differ by 11.5 and 10.5% respectively from that of the 30bar BPST model. A similar comparison under the efficient mill condition (Figure 39) shows that the specific operating cost of the Partial Fast Pyrolysis and the Pure Fast Pyrolysis BMECP models are 65.1% and 145.3% respectively higher than the specific operating cost of the 30bar BPST whereas their total operating cost are 12.0% and 10.8% respectively higher. The high energy required to run the pyrolysis process results in low net electricity output from these pyrolysis based BMECP plants thus leading to the high specific operating cost. However, these two pyrolysis based BMECP plants in addition to electricity produce bio-oil and biochar which are not included in specific operating cost estimation. The specific operating cost of these BMECP models are thus expected to reduce when these by-products (bio-oil and biochar) are converted to electricity.

5.5.3 Sales Revenue

The total sales revenue (inclusive of main products revenue and by-products revenue) for the combustion based and the pyrolysis based BMECP models are shown in Figure 40 for both sugar mill conditions. It is seen from Figure 40 that sales revenue from the combustion based CEST systems are higher than that of the BPST system irrespective of the mill condition, with revenues increasing as CEST system operating pressure increases as a result of the increase in electricity generation with increasing pressure. Total sales revenue range from a low of $50.40 million (30bar BPST) to a high of $70.62 million (82bar CEST) under the efficient mill condition and from a low of $42.13 million (30bar BPST) to a high of $67.03 million (82bar CEST) under the less efficient mill condition. Due to the fact that the combustion based BMECP models produce more electricity under the efficient condition than under the less efficient mill condition, their sales revenues are also higher under the conditions of the efficient mill as shown in Figure 40.

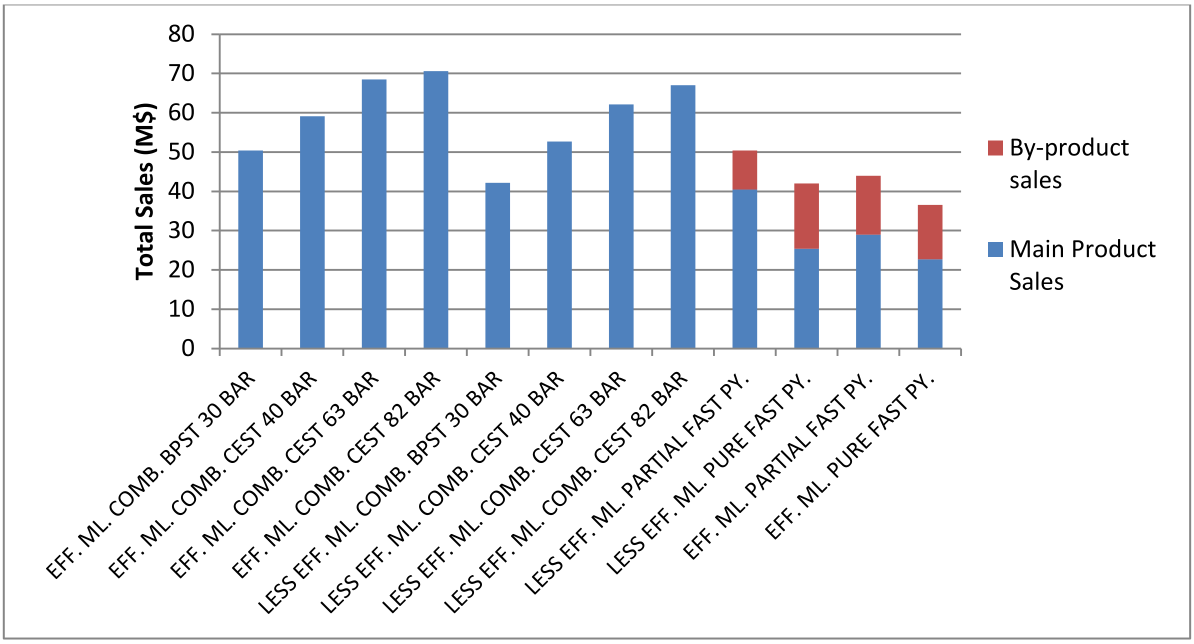

Figure 40: Total Sales Revenue for BMECP models under both efficient and less efficient mills conditions

For the pyrolysis based BMECP models the opposite is seen when total sales revenues under the efficient and the less efficient mill conditions are compared. The Partial Fast Pyrolysis BMECP and the Pure Fast Pyrolysis BMECP models generate sales revenues of $50.41 million and $41.97 million respectively under the less efficient mill which are approximately 14.8% and 15.0% higher than the sales revenues they generate when implemented in an efficient mill respectively. The reason for this is that the pyrolysis based BMECP models produce more electricity under the less efficient mill condition than under the efficient mill condition.

A breakdown of the total sales revenue of both Partial Fast Pyrolysis and Pure Fast Pyrolysis BMECP models indicates the revenue from main product (electricity) sales makes the dominant contribution as shown in Figure 40. This is because only a small fraction (depending on mill efficiency) of the bagasse feed fed to the Partial Fast Pyrolysis BMECP is used for the production of bio-oil and biochar as by-products. Also all the biochar and about half of the bio-oil produced by the pure fast pyrolysis BMECP model is combusted for electricity and to generate the quantity of steam required by the sugar mill thus leaving little by-products for sale. A possible solution to averting this problem is to consider the use of SCAR in the combustor for steam and electricity production and using all available bagasse for bio-oil and biochar production or combusting available bagasse for steam and electricity and producing bio-oil and biochar from SCAR. In this way more by-products would be produced by the Partial Fast Pyrolysis BMECP model while all bio-oil and biochar produced by the

Pure Fast Pyrolysis model would also be available for sale directly or converted to electricity for sale to increase total sales revenue. However, SCAR has got a high ash content compared to bagasse (see Table 9 in chapter 2) and might result in the production of low quality by-products (bio-oil and biochar with high ash content) or fouling of heat transfer surfaces of the biomass combustor/boiler (Morris, 2001; Pippo et al., 2007) leading to increased maintenance cost and operating charges. Hence a trade-off between sales revenue and operating cost needs to be established when considering the use of SCAR in the two pyrolysis based BMECP models.

5.5.4 Profitability Indicators/Analysis

The economic viability of the BMECP models simulated in this study were compared based on NPV, IRR, PO and PI and the results are shown in Table 23 for both sugar mill scenarios.

Table 23: Profitability Indicators for BMECP Models

	Profitability Indicators			
	NPV ($)	IRR (%)	PO (year)	PI
EFFICIENT MILL				
Combustion				
30bar BPST	256169000	36.31	5.01	1.49
40bar CEST	316818000	35.39	5.14	1.53
63bar CEST	389971000	34.51	5.26	1.58
82bar CEST	364641000	29.98	6.08	1.50
Pure Fast Pyrolysis	25938800	13.80	16.32	0.00
Partial Fast Pyrolysis	127440000	20.76	9.10	1.23
LESS EFFICIENT MILL				
Combustion				
30bar BPST	164719000	28.38	6.46	1.34
40bar CEST	261351000	31.64	5.89	1.47
63bar CEST	329957000	30.80	5.91	1.49
82bar CEST	353490000	29.62	6.16	1.50
Pure Fast Pyrolysis	78623600	15.30	12.34	0.00
Partial Fast Pyrolysis	197748000	25.42	7.26	1.34

Analysis of the results in Table 23 under the efficient mill shows that all the BMECP models give a positive NPV at the end of the respective projects life, giving indication that all the BMECP models simulated must be profitable based on the economic assumptions used in this study. However, the 63bar CEST combustion based model records the highest NPV value of about $390 million and hence seems to be the most favourable option among the various BMECP models based on NPV analysis. The least favourable option based on NPV is the Pure Fast Pyrolysis model, having NPV of $26 million at the end of the project life. The ranking order of profitability based on IRR and PO approaches places the 30bar BPST combustion based models as the most profitable option followed by the 40bar, 63bar and 82bar CEST combustion based models, the Partial Fast Pyrolysis model and the least being the Pure Fast Pyrolysis BMECP model with IRR of 13.8% (only 3.8% points higher than the 10% discount/interest rate used in the analysis) and PO of 16.32 years out of a total project life of 20 years. Based on PI approach, the 63bar CEST system has the highest value of 1.58, making it the most profitable while the Pure Fast Pyrolysis models is the least profitable with 0.00 PI. Therefore from an economic point of view, the Pure Fast Pyrolysis BMECP is the option with least economic viability under the efficient mill conditions as it records the lowest NPV, IRR, PI and the highest PO values. The IRR of the 63bar CEST model when compared to that of the 30bar BPST, differ by only 1.8% points while PO values are also similar; 5.26 versus 5.01 years. Thus given the fact that the 63bar CEST BMECP has the highest NPV and PI, it seems to be most economic viable BMECP technology for an efficient mill.

By considering the results under the less efficient mill scenario, the 82bar CEST combustion based BMECP stands out to be the most favourable option based on NPV approach. The IRR and the PO approaches however both favour the 40bar CEST BMECP while analysis based on PI tends to favour both the 63bar CEST and the 82bar CEST BMECP models. Again the Pure Fast Pyrolysis BMECP model appears to be the least economic viable BMECP option recording lowest values of about $79 million, 15.3% and 0.00 for NPV, IRR and PI respectively and the highest PO of 12.34 years.

Comparison of the profitability indicators of the two pyrolysis based BMECP models to those of the combustion based models shows the pyrolysis based models are less economical than the combustion based models. The Partial Fast Pyrolysis BMECP model tends to perform better than the Pure Fast Pyrolysis model under both mill conditions. It shows even much potential under the less efficient mill and with some process refinement within the sugar mill to cut down energy usage, could compete with the combustion based models. The Partial Fast Pyrolysis BMECP model is therefore the most economic viable option to consider as far as pyrolysis implementation in the sugar mill for steam, electricity and pyrolysis energy products production is concern.

5.5.5 Sensitivity Analysis

The profitability indicators of the various BMECP models analysed in section 5.4.4 above depend on parameter choices such as raw material (bagasse) price, electricity sales price and discount/interest rate. A sensitivity analysis was thus used to study the effects of changes in these parameters on the overall economics of the various BMECP models. For bagasse and electricity prices, a ±30% change with respect to initial values (see Table 20) was assumed and used to estimate the corresponding changes in NPV and IRR values. The range for discount/interest rate used in the sensitivity analysis was from 5 to 30%. The results of the analysis are presented in Figure 41 and Figure 42 and Table 24 and Table 25. Figure 41 shows the percentage variation in NPV values of the various models to changes in bagasse price and Figure 41 shows the percentage variation in NPV values to changes in electricity price. Changes in NPV values to changes in discount rates are outlined in Tables 24 and 25. Detail of the sensitivity analysis (including changes in IRR values to changes in the above mentioned parameters) is given in Appendix D3.

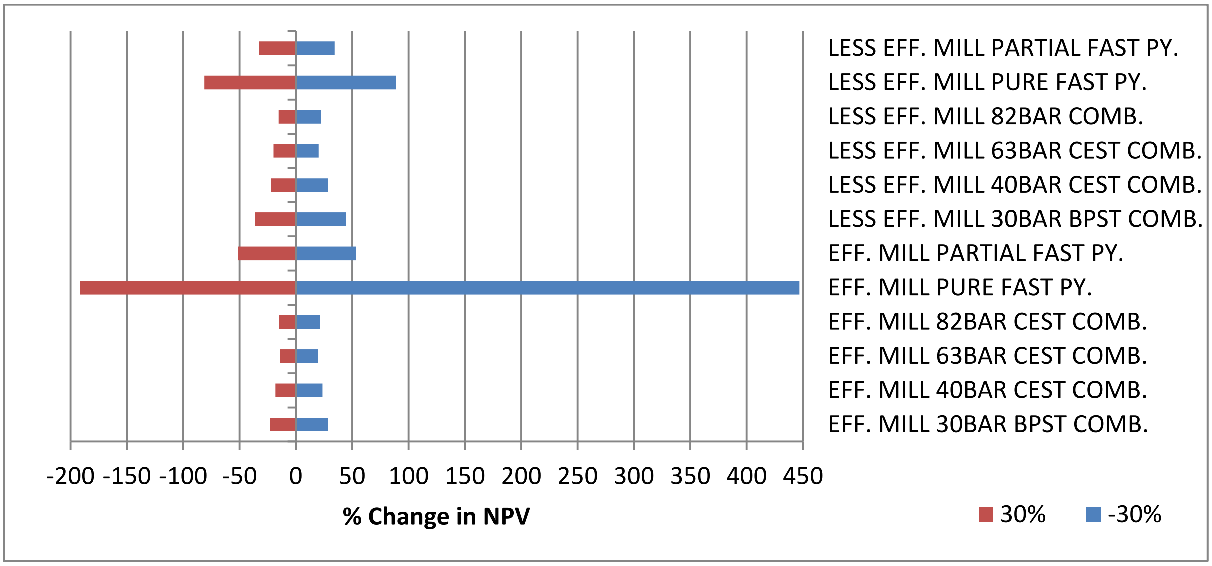

Figure 41: Percentage Variation in NPV Values of BMECP Models to Changes in Bagasse Price

From Figure 41, the Pure Fast Pyrolysis BMECP model is seen to be the model most strongly affected by changes in bagasse cost irrespective of the mill condition due to its associated high capital cost and low sales revenue it generates. Its NPV changes by as high as -191.61/+446.86% for a ±30% change in bagasse price under the efficient mill condition and -81.21/+88.64% under the less efficient mill. The least affected models are the 63bar CEST and 82bar CEST combustion based models. When bagasse price increases by 30%, the NPV of the 63bar CEST system reduces by 14.32 and 19.94% under the efficient and less efficient mill conditions respectively while the NPV of the 82bar CEST reduces by 14.85 and 15.32% under the efficient and the less efficient mills respectively. The corresponding changes to 30% decrease in bagasse price are 19.58 and 20.14% for the 63bar CEST and 21.42 and 22.08% for the 82bar CEST respectively under the efficient and the less efficient mill conditions. Despite the high TCIs associated with these BMECP models, they produce enough electricity thus generating high sales revenues to counter the effect of rise in bagasse price.

The effect of electricity price change on NPV (Figure 42) shows a similar trend to the one observed with regards to bagasse price change. Again the Pure Fast Pyrolysis BMECP model is seen to be the one most affected by both positive and negative changes in electricity price while the 63 bar CEST is the least affected, hence making the Pure Fast Pyrolysis model more susceptible to market fluctuations.

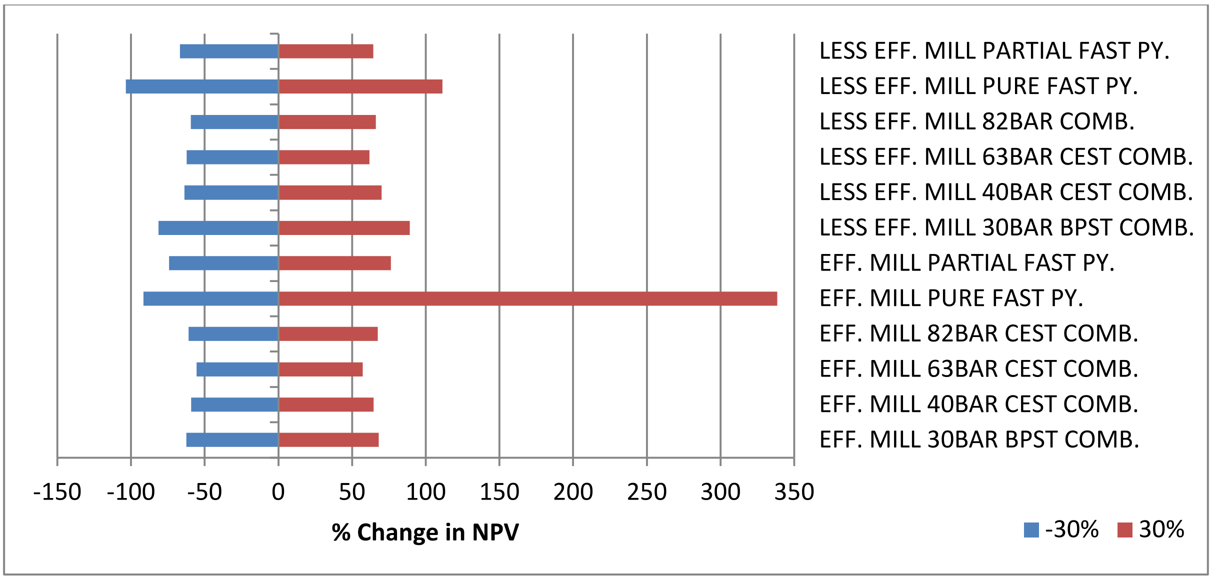

Figure 42: Percentage Variation in NPV Values of BMECP Models to Changes in Electricity Price

In Table 24, it is seen that the combustion based BMECP models still yield positive NPVs even when interest rate is increased to a high of 30% while the pyrolysis based models yield negative NPVs, with the Pure Fast Pyrolysis model yielding a negative NPV even for just 5% points increase in interest rate from 10 to 15%. It could be noted from Table 25 that while the pyrolysis based models continue to yield negative NPV values at higher interest rate at the end of the project life, the 40bar CEST and the 63bar CEST combustion models still produce positive NPV values. However, the 82bar CEST and the 30bar BPST models also yield negative NPVs, indicating the need for improvement in mill efficiency.

Table 24: Effect of Interest Rate on NPV of BMECP Models (Efficient Mill)

INTEREST RATE (%)	NPV (M$)					
	COMB. BPST 30 BAR	COMB. CEST 40 BAR	COMB. CEST 63 BAR	COMB. CEST 82 BAR	PURE FAST PYROLYSIS	PARTIAL FAST PYROLYSIS
5	497.095	617.56	764.038	741.46	244.627	299.8
10	256.169	316.818	389.971	364.641	25.392	127.44
15	144.579	177.07	216.032	192.775	-19.165	45.353
20	79.948	96.42	115.85	93.351	-53.961	1.067
30	19.31	20.906	22.195	1.366	-78.91	-37.844

Table 25: Effect of Interest Rate on NPV of BMECP Models (Less Efficient Mill)

INEREST RATE (%)	NPV (M$)					
	COMB. BPST 30 BAR	COMB. CEST 40 BAR	COMB. CEST 63 BAR	COMB. CEST 82 BAR	PURE FAST PYROLYSIS	PARTIAL FAST PYROLYSIS
5	341.899	505.928	645.766	721.624	244.277	419.898
10	164.719	261.325	329.957	353.49	78.623	197.748
15	86.188	150.079	171.16	186.01	3.214	90.06
20	40.006	84.677	84.999	89.036	-37.203	31.43
30	-2.538	23.178	5.109	-0.6	-70.943	-21.499

The break-even prices of electricity, bagasse, bio-oil and biochar at which the projects capital cost would be fully recovered at the end of the project economic life without any additional profits were also determined for each BMECP model as part of the sensitivity analysis and the results are shown in Table 26.

Table 26: Break-Even Prices of Raw Material, Products and By-products

	BREAK-EVEN PRICES			
	ELECTRICITY. ($/kWh)	BAGASSE ($/DRY TON)	BIO-OIL ($/TON)	BIOCHAR ($/TON)
EFFICIENT MILL				
Combustion				
30bar BPST	0.182	122.00		n/a
40bar CEST	0.173	137.59		
63bar CEST	0.162	156.49		
82bar CEST	0.178	150.44		
Pure Fast Pyrolysis	0.235	68.99	137.76	n/a
Partial Fast Pyrolysis	0.205	88.22	14.50	No effect
LESS EFFICIENT MILL				
Combustion				
30bar BPST	0.217	98.9		n/a
40bar CEST	0.184			
63bar CEST	0.178	138.82		
82bar CEST	0.174	147.66		
Pure Fast Pyrolysis	0.220	76.48	94.21	n/a
Partial Fast Pyrolysis	0.189	105.82	No effect	No effect

It is seen from Table 26 that the break-even price for the pyrolysis based models with regards to electricity price are much closer to the base value ($0.248/kWh) than for the combustion based models. For example, while electricity can be sold for a low of $0.162/kWh for the 63bar CEST to still break-even, the Pure Fast Pyrolysis BMECP models can only break-even at electricity price of \$0.235/kWh under the efficient mill condition. With regards to break-even bagasse price, Table 26 shows that the combustion based models can break-even at bagasse prices that are about two to three times the base case bagasse price of \$56/dry ton under the efficient mill condition and about two times the base case value under the less efficient mill condition. The highest break-even bagasse price of \$156.49/dry ton is recorded by the 63bar CEST BMECP model under the efficient mill condition while the 82bar CEST BMECP model records the highest break-even bagasse price of \$147.66/dry ton under the less efficient mill. The break-even bagasse price for the pyrolysis based models shown in Table 26 are slightly higher than the base price but are significantly lower than those of the combustion based models. However, there is an exception under the less efficient mill scenarios where the Partial Fast Pyrolysis BMECP model achieves a break-even bagasse price of \$105.82/dry ton (about twice the base value), which is about \$7 higher than the break-even bagasse price of the 30bar BPST system. This is because the Partial Fast Pyrolysis BMECP model under the condition of the less efficient mill generates almost the same amount of electricity as the 30bar BPST model; 20.78 vs. 21.40 MW (see Table 16 and also Table D2.2 in Appendix D) in addition to pyrolysis products. Additional revenue generated from the sale of these by-products therefore gives the Partial Fast Pyrolysis model overall total sales revenue higher than that of the 30bar BPST BMECP model (refer to Figure 40 and Table D2.2), thus giving it a higher break-even bagasse price. This also explains why the break-even electricity price of the 30bar BMECP model is lower than that of the Partial Fast Pyrolysis BMECP model under conditions of the less efficient mill. This trend can also be seen in both Figure 41 and 42.

The bio-oil and biochar break-even prices could not be determined for the combustion based models because these models produce no bio-oil and biochar. Also the break-even biochar price could not be determined for the Pure Fast pyrolysis BMECP model as all the biochar produced by this BMECP model is utilised as fuel in the combustor for process energy generation as explained earlier. It is seen from Table 26 that the profitability of the Pure Fast Pyrolysis BMECP model relies very much on the sales revenue from bio-oil as this model only break-even at bio-oil prices which are only 18.96 and 44.58% less than the base value of \$170/ton under the efficient and less efficient mills conditions respectively. This might be caused by the low electricity sales revenue it generates due to its low surplus electricity production rate. The analysis also show that even though the Partial Fast Pyrolysis BMECP model produces bio-oil and biochar as by-products, the sales revenue from these by-products have virtually no effect on the profitability of this model. This is because in the current study not much by-products are produced by this BMECP model as a result of the bypass of a significant quantity of bagasse to the combustor for steam and electricity generation in order to satisfy the energy demands of both the sugar mill and the pyrolysis process itself. However, in a situation where another biomass source such as SCAR is used to replace the bagasse in the combustor thus allowing enough bagasse to be converted to pyrolysis products, this model would produce significant amount of bio-oil and biochar in addition to electricity which would impact much on its profitability whether the by-products are sold directly or converted to electricity for sale.

As stated earlier, the BMECP plant was assumed to be located close to the sugar mill receives its fuel supply (bagasse) directly from the mill and in turn provides the sugar mill with steam and electricity. This implies that under normal circumstance the bagasse used by the various BMECPs would be delivered free of charge (Zero cost to the BMECP). Also the current cost of coal derived electricity in South Africa is 65cents/kWh as opposed to the NERSA price for bagasse derived electricity used in this study for the economic analysis. These prices (zero bagasse cost and 65cent/kWh electricity price) were therefore also adopted and incorporated into the economic analysis to study their effects on the economic viability of the various BMECP models. The results for these analyses are presented in Tables 27 – 29.

In Table 27, 65cent/kWh electricity price was assumed while the bagasse price was maintained at the base price (see Table 20). As seen in Table 27 all the BMECP models are not profitable under these conditions irrespective of the sugar mill scenario. All BMECPs recorded negative NPV while their IRR and PO could not be determined. This indicates that bagasse derived electricity in South Africa can only be profitable under such conditions when electricity price is above 65 cent/kWh.

Table 27: BMECPs Response to 65cent/kWh Electricity Price and Base Case Bagasse Cost

	Profitability Indicators			
	NPV ($)	IRR (%)	PO (year)	PI
EFFICIENT MILL				
Combustion				
30bar BPST	-122576000	n/a	n/a	0.69
40bar CEST	-118904000	n/a	n/a	0.73
63bar CEST	-102822000	n/a	n/a	0.78
82bar CEST	-152737000	n/a	n/a	0.71
Pure Fast Pyrolysis	-72593400	n/a	n/a	n/a
Partial Fast Pyrolysis	-79752700	n/a	n/a	0.82
LESS EFFICIENT MILL				
Combustion				
30bar BPST	-168337000	n/a	n/a	0.58
40bar CEST	-138028000	n/a	n/a	0.67
63bar CEST	-136465000	n/a	n/a	0.71
82bar CEST	-133156000	n/a	n/a	0.73
Pure Fast Pyrolysis	-105655000	n/a	n/a	n/a
Partial Fast Pyrolysis	-93481800	n/a	n/a	0.79

At zero bagasse cost and 65 cent/kWh electricity price, the result as seen in Table 28 indicates that all the BMECPs are profitable. However, the pyrolysis based models tend to be more profitable than the two advance combustion systems (63bar CEST and 82bar CEST). This might be due to the production of high valued pyrolysis products which generate extra income in addition to the income from the sales of electricity. Though the capital cost of the pyrolysis based BMECPs are just as high as that of the two advanced combustion systems, the extra income generated compensate for the short fall on sales revenue when there is a drop in electricity price as compared to the 63bar CEST and 82bar CEST systems whose sales revenues depends entirely on electricity sales. Also it can be seen form Table 28 that both the 30bar BPST and the 40bar CEST combustion BMECPs are more economically viable than the 63bar and the 82bar CEST systems at conditions of zero bagasse cost and 65cent/kWh electricity price. This is due the high capital cost associated with the 63bar and the 82bar CEST systems as compared to that of 30bar BPST and the 40bar CEST systems.

Table 28: BMECPs Response to Zero Bagasse Cost and 65cent/kWh Electricity Price

	Profitability Indicators			
	NPV ($)	IRR (%)	PO (year)	PI
EFFICIENT MILL				
Combustion				
30bar BPST	120947000	24.90	7.41	1.75
40bar CEST	119599000	21.73	8.62	1.57
63bar CEST	126284000	19.82	9.57	1.49
82bar CEST	85794600	15.94	12.26	1.27
Pure Fast Pyrolysis	151716000	21.85	8.82	1.53
Partial Fast Pyrolysis	143734000	22.33	8.34	1.58
LESS EFFICIENT MILL				
Combustion				
30bar BPST	88190600	21.55	8.70	1.60
40bar CEST	109885000	23.98	7.73	1.60
63bar CEST	101942000	18.16	10.57	1.41
82bar CEST	100104000	16.88	11.50	1.36
Pure Fast Pyrolysis	122234000	19.15	10.06	1.48
Partial Fast Pyrolysis	133105000	21.45	8.73	1.54

In Table 29 a bagasse cost of zero was assumed while the price of electricity was maintained at the base price (refer to Table 20). The results show a much improvement in the profitability of the BMECP models under such conditions compared to the base economic conditions used as in Table 20 as a result of the huge reduction in total operating cost due the free bagasse.

Table 29: BMECPs Response to Zero Bagasse Cost and Base Case Electricity Price

	Profitability Indicators			
	NPV ($)	IRR (%)	PO (year)	PI
EFFICIENT MILL				
Combustion				
30bar BPST	478938000	57.43	3.32	2.61
40bar CEST	483750000	58.58	3.28	2.48
63bar CEST	612740000	47.13	3.91	2.38
82bar CEST	587409000	40.67	4.47	2.17
Pure Fast Pyrolysis	248797000	29.48	6.79	1.89
Partial Fast Pyrolysis	347619000	35.54	5.07	2.07
LESS EFFICIENT MILL				
Combustion				
30bar BPST	387488000	50.08	3.71	2.48
40bar CEST	539587000	51.65	3.62	2.45
63bar CEST	542726000	43.54	4.19	2.31
82bar CEST	576259000	40.41	4.49	2.25
Pure Fast Pyrolysis	300828000	32.33	6.32	2.04
Partial Fast Pyrolysis	417927000	39.41	4.59	2.18

CHAPTER SIX

Conclusions and Recommendations

6.1 Conclusions

The objective of this study was to develop process models for the efficient utilization of sugar mill biomass (bagasse) aimed at producing energy from bagasse to run the operations of two sugar mill configurations (efficient and less efficient mill), while also generating extra energy in the form of electricity and/or high value pyrolysis products for sale. Detailed BMECP models were developed in Aspen Plus® based on Combustion and Fast Pyrolysis thermochemical process technologies. The combustion based BMECP models developed were a 30bar BPST system (simulating the base case scenario) and 40bar, 63bar, and 82bar CEST systems. Based on how bagasse is used, Pure Fast Pyrolysis and Partial Fast Pyrolysis BMECP models were developed based on the fast pyrolysis process technology. The BMECP models were then compared based on the following technical performances; steam and electricity production rates, process efficiencies and environmental impacts (based on CO_2 savings). Effect of bagasse moisture content on steam and electricity production rates of the combustion based BMECP models was also studied by varying the moisture content of the bagasse feedstock fed to these models from 46 -50% instead of the base case value of 50% used in the other BMECP models. Moreover, detailed economic analysis models based on n^{th} plant approach were developed in Aspen Process Economic Analyzer using economic data relevant to South Africa and the results analyzed to establish the profitability of the various BMECP models. The following conclusions are made based on the results of the study:

Technical Performance of BMECP models

- All BMECP models were able to generate the quantity of steam required to run the operations of the two sugar mill configurations. However, total steam production rates of the combustion based BMECP models are higher than that of the pyrolysis based BMECP models.
- Due to the need to also produce high energy value pyrolysis products (bio-oil and biochar) the Pure Fast Pyrolysis and the Partial Fast Pyrolysis BMECP models could only produce steam just sufficient to satisfy the requirement of the mill.
- Steam production rate of the combustion based models decreases with increase in boiler operating pressure due to the increase in steam superheat conditions.
- Effect of bagasse moisture shows that steam production rate increases with decreasing bagasse moisture content. As moisture content decreases the heating value of bagasse increases thereby increasing the quantity of heat available from the biomass combustor for

steam production and hence the increase in steam production rate with decreasing moisture content.

- The combustion based models produces higher electricity production rates than the pyrolysis based models because of their high total steam production rates. Electricity production was found to increase gradually as operating pressure increases due the gradual increase in the efficiency of conversion of heat energy to electricity with increasing pressure. Electricity production was also found to increase with decreasing bagasse moisture content as a result of the increase in steam production rate.
- At the same input feed rate, the electricity production rate of the CEST combustion systems were found to be significantly higher than that of the BPST system under both mill configurations.
- Due to the high energy required for pyrolysis, the electricity production rates of the pyrolysis based BMECP models are lower than that of the combustion based models. However, the Partial Fast Pyrolysis BMECP model performed well compared to the 30bar BPST system at lower mill efficiency.
- Electrical efficiencies of the combustion based BMECP models were estimated to be higher than those of the pyrolysis based BMECP models because of their high electricity production rate. However, due to the high energy value pyrolysis products produced by the pyrolysis based BMECP models, overall process efficiencies were found to be very similar for both the pyrolysis based models and the combustion based models.
- Electricity production rates and hence electricity efficiency for the combustion based BMECP models were found to be higher under the efficient mill than under the less efficient mill. Due to the low steam requirement of the efficient mill, more surplus steam (after the extraction of mill steam requirement) is made available for further conversion to electricity than is left under the less efficient mill.
- The quantity of final pyrolysis products generated by both the Pure Fast Pyrolysis and the Partial Fast Pyrolysis BMECP models were found to be limited by sugar mill efficiency. To satisfy the energy requirement of the efficient mill and the less efficient mill, 46.5 and 61% respectively of total bio-oil in addition to all biochar produced by the Pure Fast Pyrolysis BMECP model is consumed by the process. Similarly 70 and 80% of the total bagasse feed input to the Partial Fast Pyrolysis BMECP model is combusted to supply the energy required by the efficient mill and the less efficient mill respectively thus allowing only a small quantity of bagasse for conversion to pyrolysis products.
- The use of pyrolysis products for electricity and steam production was found to impact negatively on the overall efficiency (based on final pyrolysis products) of the pyrolysis process of the Pure Fast Pyrolysis BMECP model. Efficiency drops to a low of 36.67% under

the less efficient mill and 50.29% under the efficient mill compared to the 70% obtained from the model of Leibbrandt (2010) which had no external energy duty.

- The pyrolysis based BMECP models proved to contribute more towards CO_2 savings than the combustion based BMECP models.

Economic Performance

- Comparison based on TCI shows that the 82bar CEST combustion system is the most expensive BMECP model to implement having investment cost of about $140 million. This is followed by the Pure Fast Pyrolysis model. The Partial Fast pyrolysis BMECP model cost higher than the 30bar BPST and the 40bar BMECP models but cost similar to the 63bar BMECP model.
- Total operating cost and specific operating cost were found to be higher for the pyrolysis based models than for the combustion based models. Although the 82bar CEST and the 63bar CEST BMECP models have higher TCI and total operating cost these models have the lowest specific operating; 0.75 and 0.8 $/MW respectively under the less efficient mill and 0.8 and 0.74 $/MW respectively under the efficient mill compared to the highest specific operating cost of 1.52 and 2.11 $/MW for the Pure Fast Pyrolysis model under the less efficient and efficient mill respectively.
- From an economic point of view, biomass combustion based on 63bar CEST BMECP proved to be the most economically viable option. The estimated 1st order TCI for this model is about $116 million yielding NPV of $390 million at the end of the plant economic life. The estimated IRR is 34.5% with 5.26 year payout period.
- The least profitable option is the Pure Fast Pyrolysis BMECP due to its associated high TCI and high operating cost and low electricity production rate.
- Sensitivity analysis of the BMECP showed the Pure Fast Pyrolysis BMECP is the most sensitive to changes in bagasse and electricity prices. The 63bar CEST BMECP model responds the least to variations in bagasse and electricity price and have the best break-even prices of $0.162/kW and $156.49/dry ton for electricity and bagasse respectively.

6.2 Recommendations

It is recommended that a comprehensive LCA other than what was used in this study must be conducted. This would ensure that the full economic impacts associated with the various BMECP models are accounted for right from bagasse collection through to the end use of electricity and /or bio-oil and biochar products produced from the BMECP plants.

Due to the low electricity and pyrolysis products generated by the pyrolysis based models, it is recommended that the use of SCAR to produce the required energy of the sugar mill be considered. This would ensure that enough bagasse is made available for conversion to pyrolysis products which in turn would increase the profitability of these BMECP models. Even though SCAR have high ash content and might cause corrosion and fouling problems on the heat transfer surface of the biomass combustor/boiler, this problem could be reduced by pre-treating SCAR with low concentration HCl solution to reduce the ash content (Pippo et al., 2007).

It is also recommended that BMECP models based on gasification thermochemical process for the conversion of bagasse to energy and energy products be developed and integrated into the sugar mill to assess the effect of gasification on the technical and economic performance of the sugar mill.

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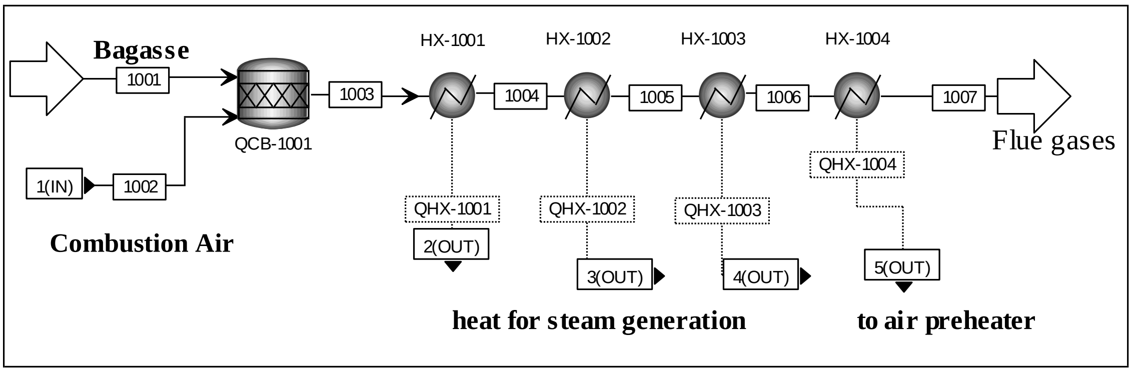
APPENDIX A – PROCESS FLOW DIAGRAMS

PFDs and Mass and Energy balances data are presented in this section for the BMECP models developed in Aspen Plus®. Due to the large number of models simulated, PFD and data are presented in this thesis only for BMECP models modelled based on the energy requirement of the efficient mill. Data for the combustion based BMECP models are also only presented for those modelled based on the 50% bagasse feed moisture content. It must however be noted that the complete set of PFDs and Mass and Energy Balances for all the models are contained on the CD attached to this thesis. To help reduce the sizes of the Mass and Energy balance table, some duplicated streams and streams with zero flow rates at certain process flow areas are omitted.

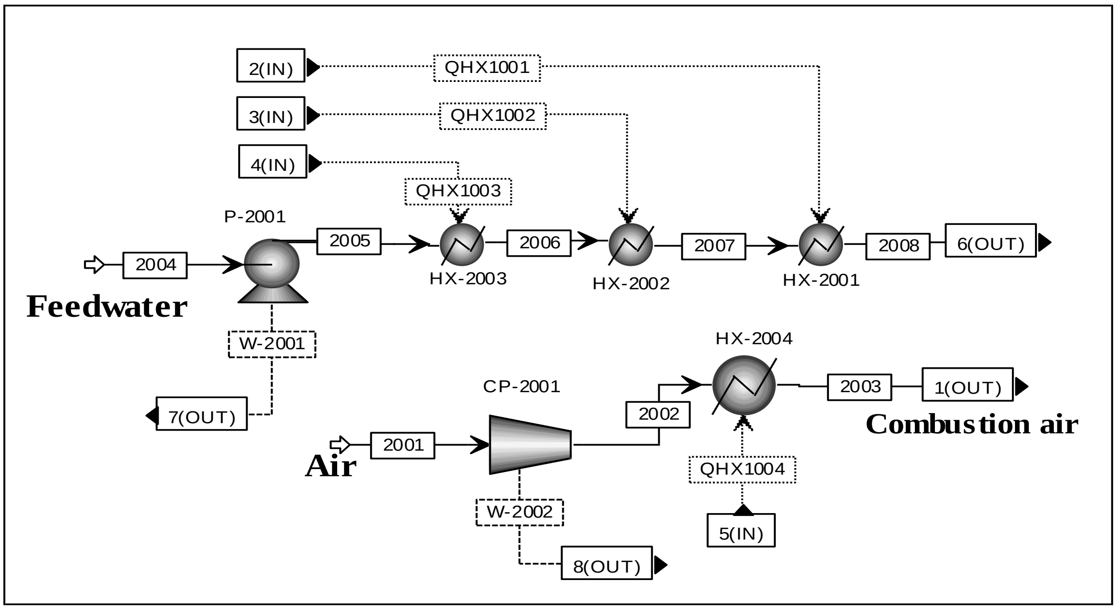
A1.1 PFD and Mass and Energy Balances for 63bar CEST BMECP (50% moisture content – efficient mill)

AREA 1000



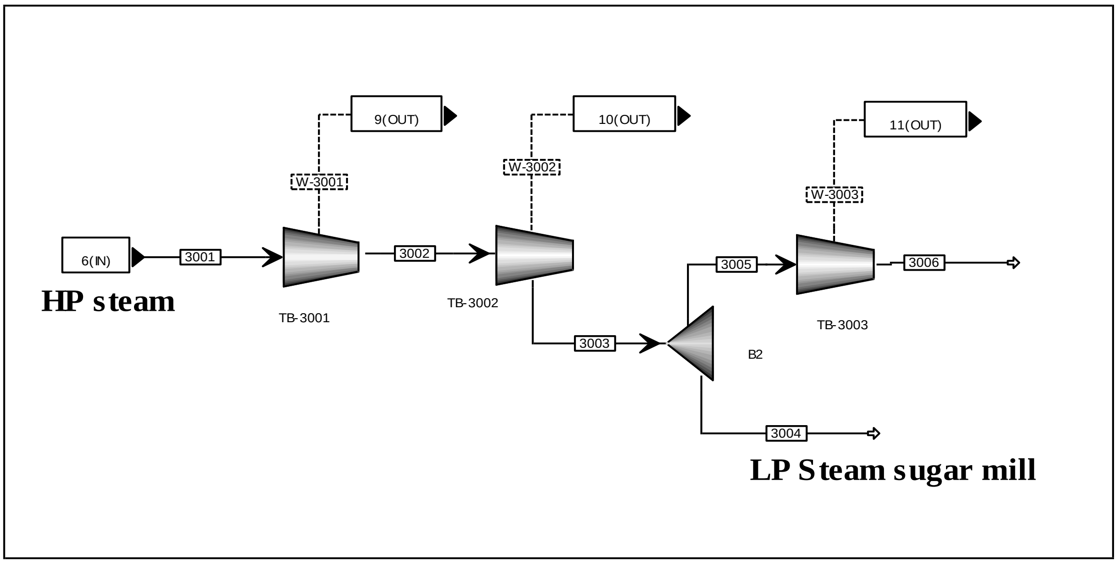
	1001	1002	1003	1004	1005	1006	1007
Mass Flow, tons/hr							
LIGNIN	10.33	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	16.44	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	3.04	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	40.50	0.00	65.26	65.26	65.26	65.26	65.26
ASH	1.46	0.00	1.46	1.46	1.46	1.46	1.46
GALACTAN	0.32	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.08	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	8.10	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.73	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	89.22	32.38	32.38	32.38	32.38	32.38
N_2	0.00	293.82	293.82	293.82	293.82	293.82	293.82
CO_2	0.00	0.00	70.59	70.59	70.59	70.59	70.59
NO_2	0.00	0.00	0.52	0.52	0.52	0.52	0.52
SO_2	0.00	0.00	0.01	0.01	0.01	0.01	0.01
SO_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	81.00	383.04	464.04	464.04	464.04	464.04	464.04
Temperature, ^{o}C	25.00	250.00	1091.77	800.00	500.00	300.90	140.00
Pressure, bar	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Vapour Fraction	0.00	1.00	1.00	1.00	1.00	1.00	1.00
Enthalpy, MJ/hr	-890420.00	79977.62	-810440.00	-980520.00	-1145400.00	-1248100.00	-1327200.00

AREA 2000



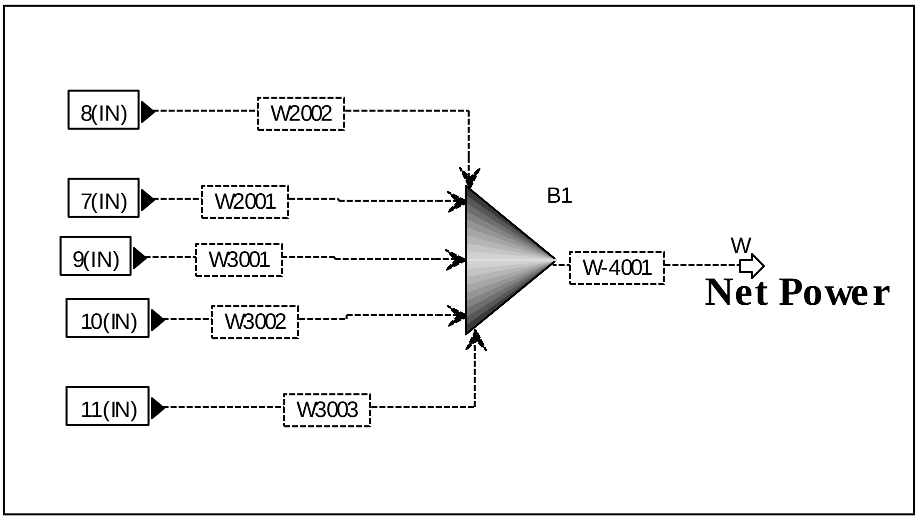
	2001	2002	2003	2004	2005	2006	2007	2008
Mass Flow, tons/hr								
LIGNIN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	0.00	0.00	153.15	153.15	153.15	153.15	153.15
ASH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
GALACTAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	89.22	89.22	89.22	0.00	0.00	0.00	0.00	0.00
N_2	293.82	293.82	293.82	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	383.04	383.04	383.04	153.15	153.15	153.15	153.15	153.15
Temperature, ^{o}C	25.00	27.39	250.00	80.00	80.84	222.57	278.79	500.00
Pressure, bar	1.00	1.02	1.00	1.30	30.00	63.00	63.00	63.00
Vapour Fraction	1.00	1.00	1.00	0.00	0.00	0.00	0.50	1.00
Enthalpy, MJ/hr	0.00	840.84	79977.62	-2172100.00	-2171600.00	-2068900.00	-1904100.00	-1734000.00

AREA 3000



	3001	3002	3003	3004	3005	3006
Mass Flow, tons/hr						
LIGNIN	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	153.15	153.15	153.15	120.00	33.15	33.15
ASH	0.00	0.00	0.00	0.00	0.00	0.00
GALACTAN	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
SO_3	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow tons/hr	153.15	153.15	153.15	120.00	33.15	33.15
Temperature, ^oC	500.00	352.31	128.89	128.89	128.89	60.06
Pressure, bar	63.00	20.00	2.00	2.00	2.00	0.20
Vapour Fraction	1.00	1.00	1.00	1.00	1.00	0.92
Enthalpy, MJ/hr	-1734000.00	-1776700.00	-1837700.00	-1440000.00	-397740.00	-407100.00

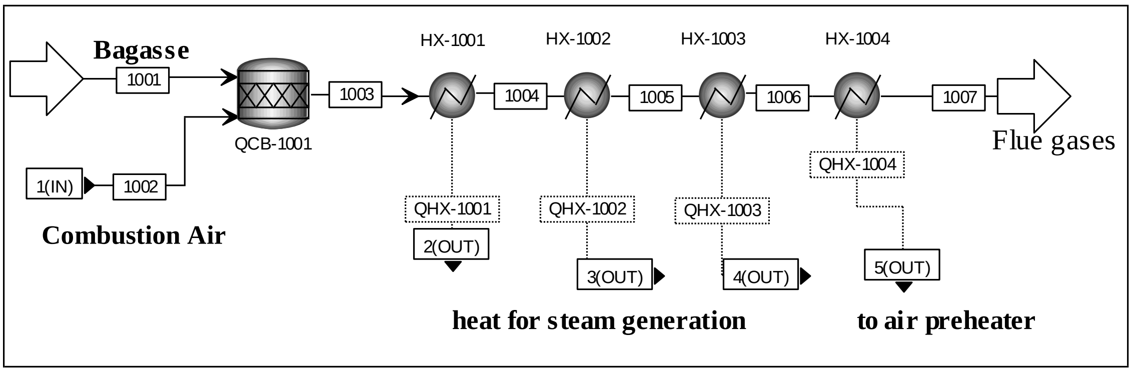
AREA 4000



	W-4001	W2001	W2002	W3001	W3002	W3003
POWER, MW	-31.03	0.14	0.23	-11.86	-16.94	-2.60
SPEED						

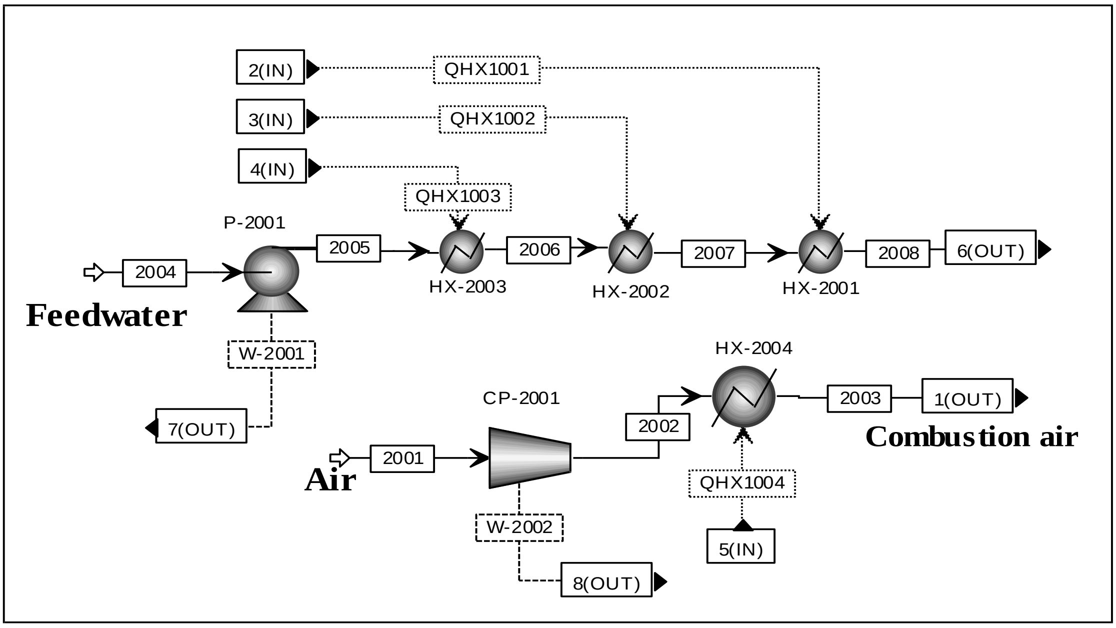
A1.2 PFD and Mass and Energy Balances for 82bar CEST BMECP (50% moisture - efficient mill)

AREA 1000



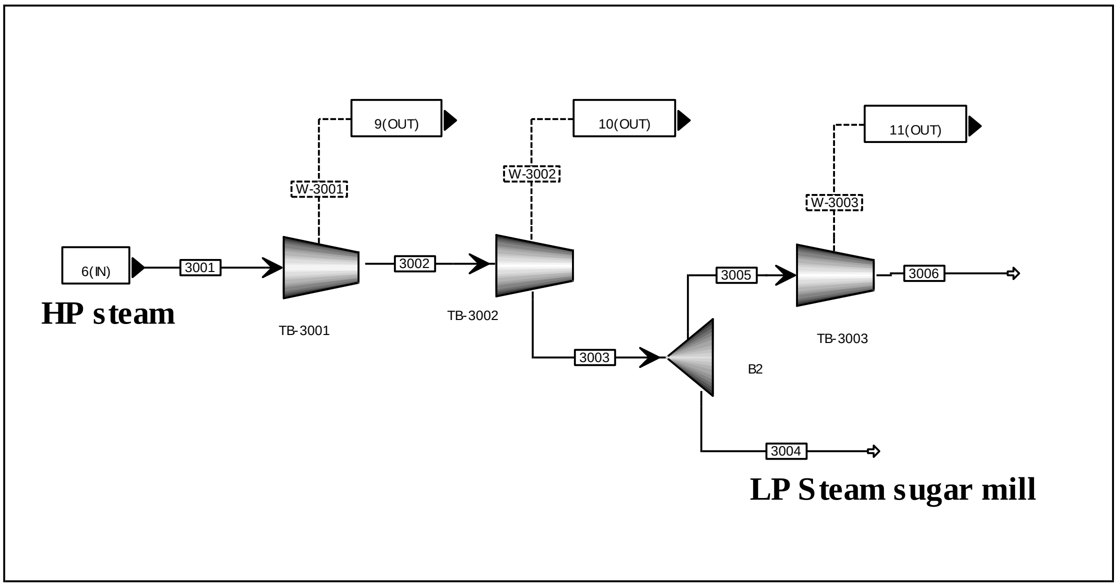
	1001	1002	1003	1004	1005	1006	1007
Mass Flow, tons/hr							
LIGNIN	10.33	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	16.44	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	3.04	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	40.50	0.00	65.26	65.26	65.26	65.26	65.26
ASH	1.46	0.00	1.46	1.46	1.46	1.46	1.46
GALACTAN	0.32	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.08	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	8.10	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.73	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	89.22	32.38	32.38	32.38	32.38	32.38
N_2	0.00	293.82	293.82	293.82	293.82	293.82	293.82
CO_2	0.00	0.00	70.59	70.59	70.59	70.59	70.59
NO_2	0.00	0.00	0.52	0.52	0.52	0.52	0.52
SO_2	0.00	0.00	0.01	0.01	0.01	0.01	0.01
SO_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	81.00	383.04	464.04	464.04	464.04	464.04	464.04
Temperature, ^oC	25.00	250.00	1091.77	800.00	500.00	300.90	140.00
Pressure, bar	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Vapour Fraction	0.00	1.00	1.00	1.00	1.00	1.00	1.00
Enthalpy, MJ/hr	-890420.00	79977.62	-810440.00	-980520.00	-1145400.00	-1248100.00	-1327200.00

AREA 2000



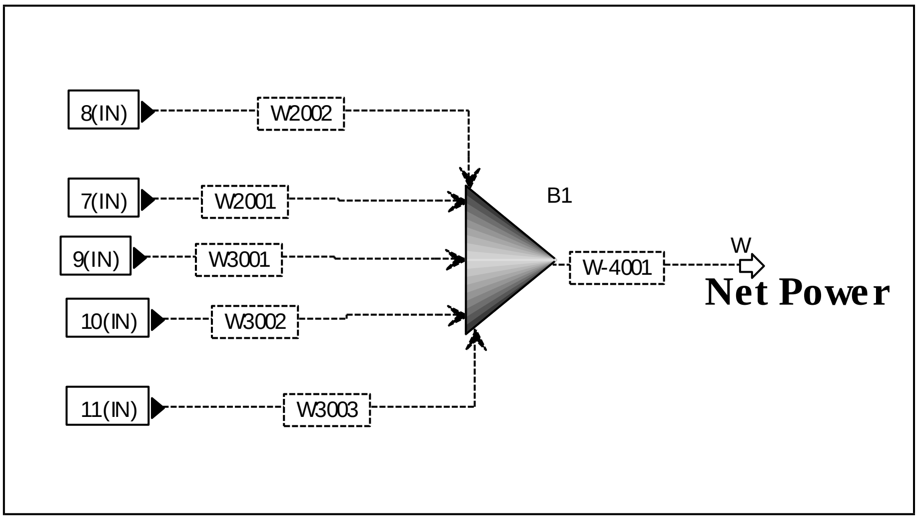
	2001	2002	2003	2004	2005	2006	2007	2008
Mass Flow, tons/hr								
LIGNIN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	0.00	0.00	150.59	150.59	150.59	150.59	150.59
ASH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
GALACTAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	89.22	89.22	89.22	0.00	0.00	0.00	0.00	0.00
N_2	293.82	293.82	293.82	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	383.04	383.04	383.04	150.59	150.59	150.59	150.59	150.59
Temperature, ^oC	25.00	27.39	250.00	80.00	80.84	224.50	296.69	525.00
Pressure, bar	1.00	1.02	1.00	1.30	30.00	82.00	82.00	82.00
Vapour Fraction	1.00	1.00	1.00	0.00	0.00	0.00	0.46	1.00
Enthalpy, MJ/hr	0.00	840.84	79977.62	-2135900.00	-2135400.00	-2032700.00	-1867800.00	-1697700.00

AREA 3000



	3001	3002	3003	3004	3005	3006
Mass Flow, tons/hr						
LIGNIN	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	150.59	150.59	150.59	120.00	30.59	30.59
ASH	0.00	0.00	0.00	0.00	0.00	0.00
GALACTAN	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
SO_3	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	150.59	150.59	150.59	120.00	30.59	30.59
Temperature, ^oC	525.00	343.01	122.05	122.05	122.05	85.96
Pressure, bar	82.00	20.00	2.00	2.00	2.00	0.60
Vapour Fraction	1.00	1.00	1.00	1.00	1.00	0.95
Enthalpy, MJ/hr	-1697700.00	-1749600.00	-1808800.00	-1441400.00	-367420.00	-372200.00

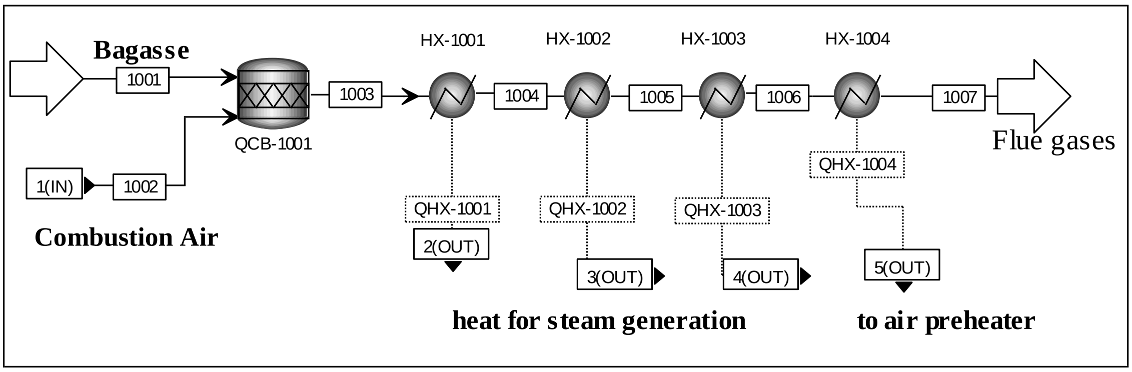
AREA 4000



	W-4001	W2001	W2002	W3001	W3002	W3003
POWER, MW	-31.81	0.14	0.23	-14.41	-16.44	-1.33
SPEED						

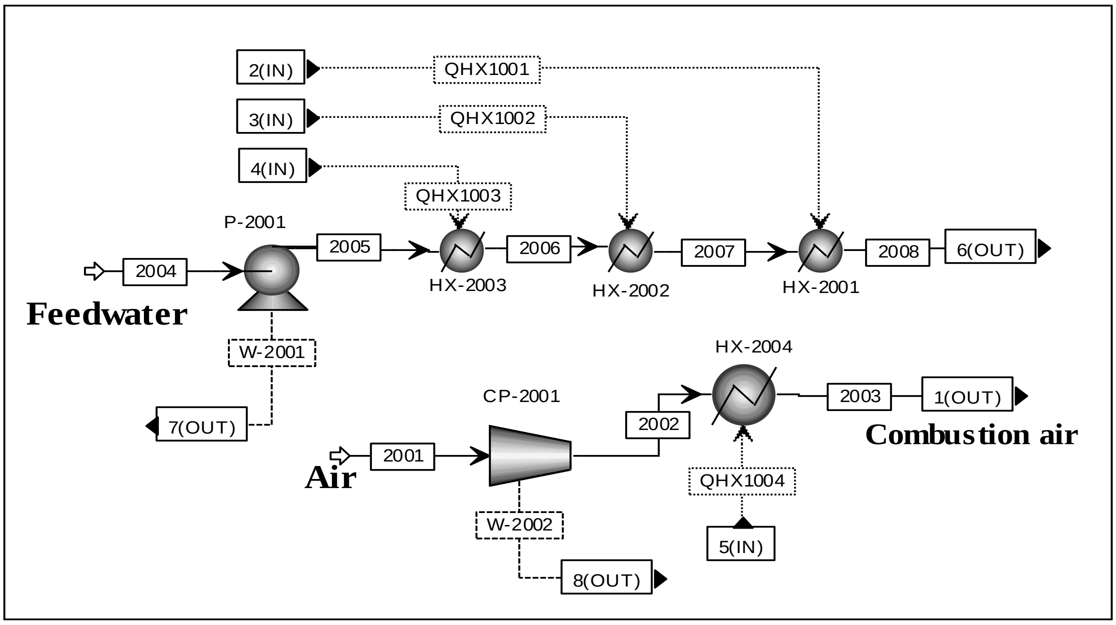
A1.3 PFD and Mass and Energy Balances for 30bar BPST BMECP (50% moisture content – efficient mill)

AREA 1000 – FEED COMBUSTION



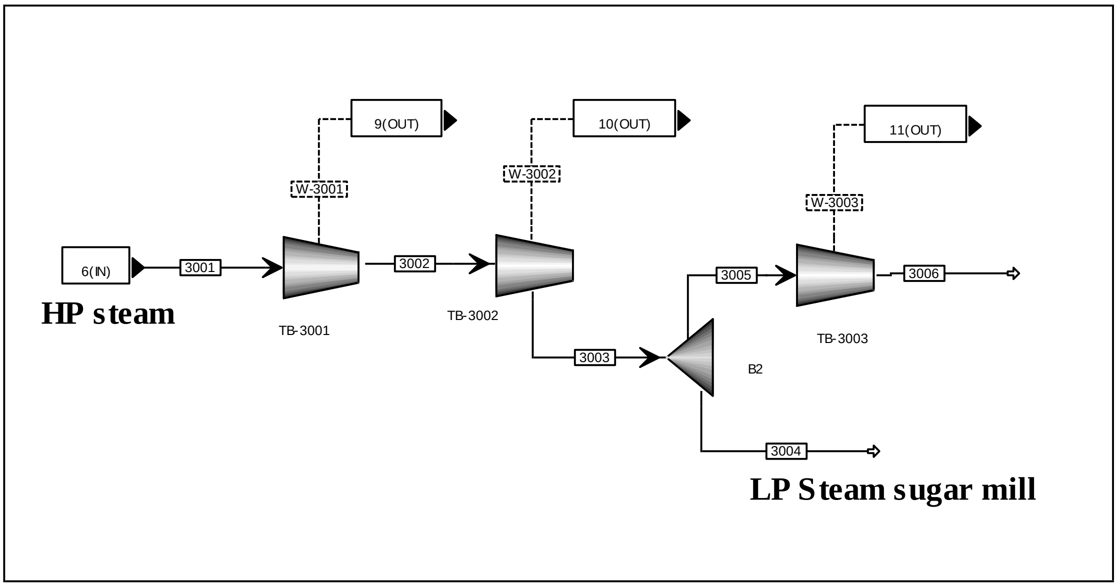
	1001	1002	1003	1004	1005	1006	1007
Mass Flow, tons/hr							
LIGNIN	10.33	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	16.44	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	3.04	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	40.50	0.00	65.26	65.26	65.26	65.26	65.26
ASH	1.46	0.00	1.46	1.46	1.46	1.46	1.46
GALACTAN	0.32	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.08	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	8.10	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.73	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	95.96	39.12	39.12	39.12	39.12	39.12
N_2	0.00	316.04	316.04	316.04	316.04	316.04	316.04
CO_2	0.00	0.00	70.59	70.59	70.59	70.59	70.59
NO_2	0.00	0.00	0.52	0.52	0.52	0.52	0.52
SO_2	0.00	0.00	0.01	0.01	0.01	0.01	0.01
SO_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	81.00	412.00	493.00	493.00	493.00	493.00	493.00
Temperature, ^oC	25.00	250.00	1051.79	800.00	500.00	303.93	140.00
Pressure, bar	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Vapour Fraction	0.00	1.00	1.00	1.00	1.00	1.00	1.00
Enthalpy, MJ/hr	-890420.00	86024.40	-804390.00	-958510.00	-1132300.00	-1239000.00	-1324100.00

AREA 2000 – HEAT RECOVERY AND STEAM PRODUCTION



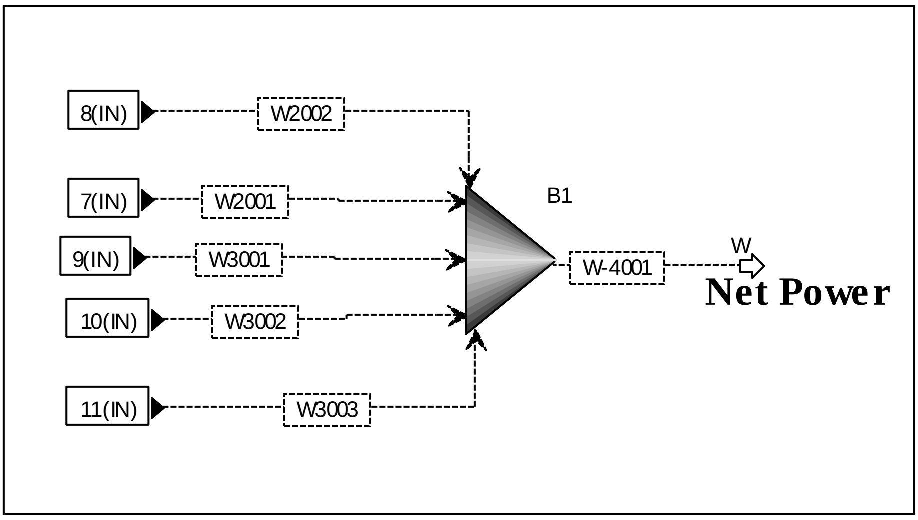
	2001	2002	2003	2004	2005	2006	2007	2008
Mass Flow, tons/hr								
LIGNIN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	0.00	0.00	160.62	160.62	160.62	160.62	160.62
ASH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
GALACTAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	95.96	95.96	95.96	0.00	0.00	0.00	0.00	0.00
N_2	316.04	316.04	316.04	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	412.00	412.00	412.00	160.62	160.62	160.62	160.62	160.62
Temperature, °C	25.00	27.39	250.00	70.00	70.85	214.85	233.94	400.00
Pressure, bar	1.00	1.02	1.00	1.30	30.00	30.00	30.00	30.00
Vapour Fraction	1.00	1.00	1.00	0.00	0.00	0.00	0.60	1.00
Enthalpy, MJ/hr	0.00	904.41	86024.40	-2284300.00	-2283800.00	-2177100.00	-2003300.00	-1849200.00

AREA 3000 – STEAM TURBINE AND POWER CYCLE



	3001	3002	3003	3004	3005
Mass Flow, tons/hr					
LIGNIN	0.00	0.00	0.00	0.00	0.00
CELLULOS	0.00	0.00	0.00	0.00	0.00
EXTRACT	0.00	0.00	0.00	0.00	0.00
H_2O	160.62	160.62	40.62	120.00	40.62
ASH	0.00	0.00	0.00	0.00	0.00
GALACTAN	0.00	0.00	0.00	0.00	0.00
MANNAN	0.00	0.00	0.00	0.00	0.00
XYLAN	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00
SO_3	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	160.62	160.62	40.62	120.00	40.62
Temperature, ^{o}C	400.00	130.15	130.15	130.15	60.06
Pressure, bar	30.00	2.00	2.00	2.00	0.20
Vapour Fraction	1.00	1.00	1.00	1.00	0.92
Enthalpy, MJ/hr	-1849200.00	-1927100.00	-487360.00	-1439700.00	-498840.00

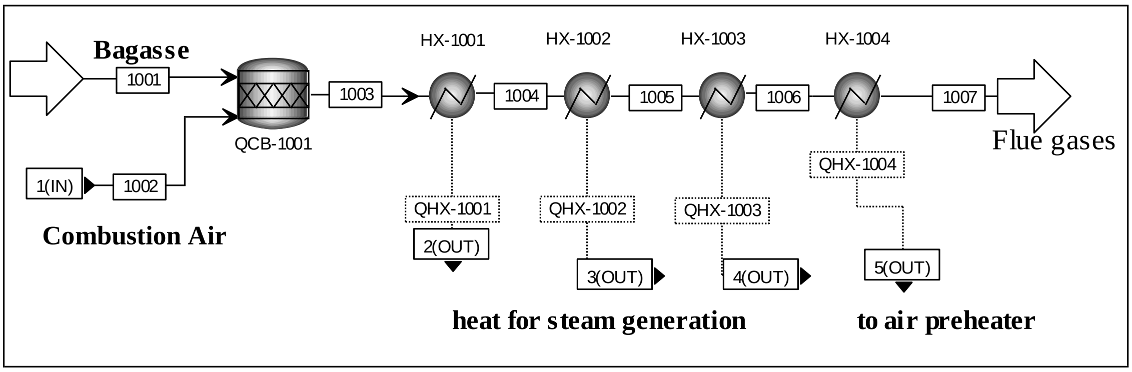
AREA 4000 – UTILITIES



	W-4001	W2001	W2002	W3001	W3002
POWER, MW	-24.42	0.14	0.25	-21.63	-3.19
SPEED					

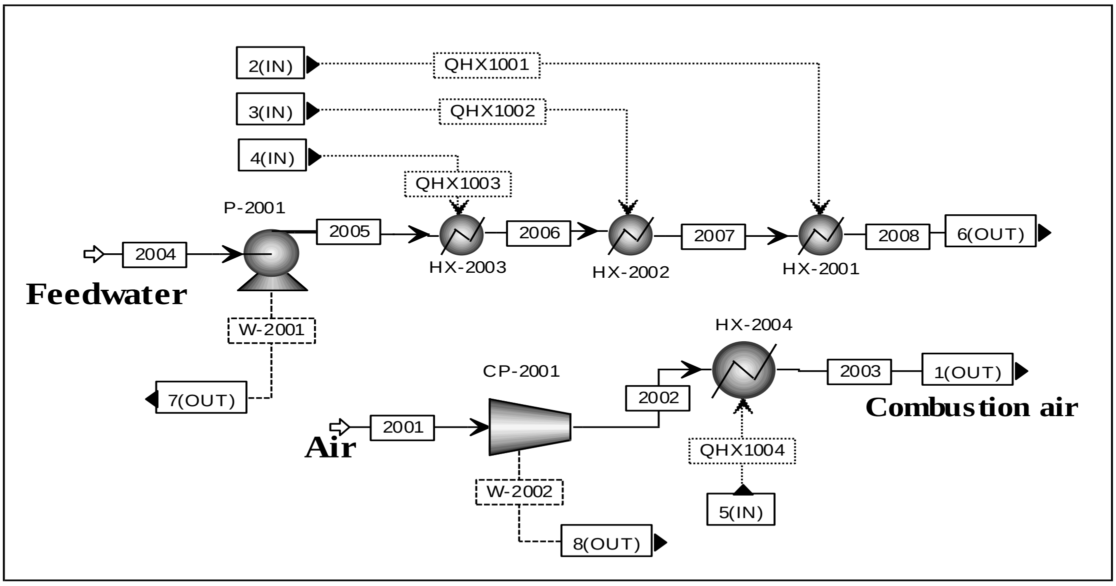
A1.4 PFD and Mass and Energy Balances for 40bar CEST BMECP (50% moisture – efficient mill)

AREA 1000



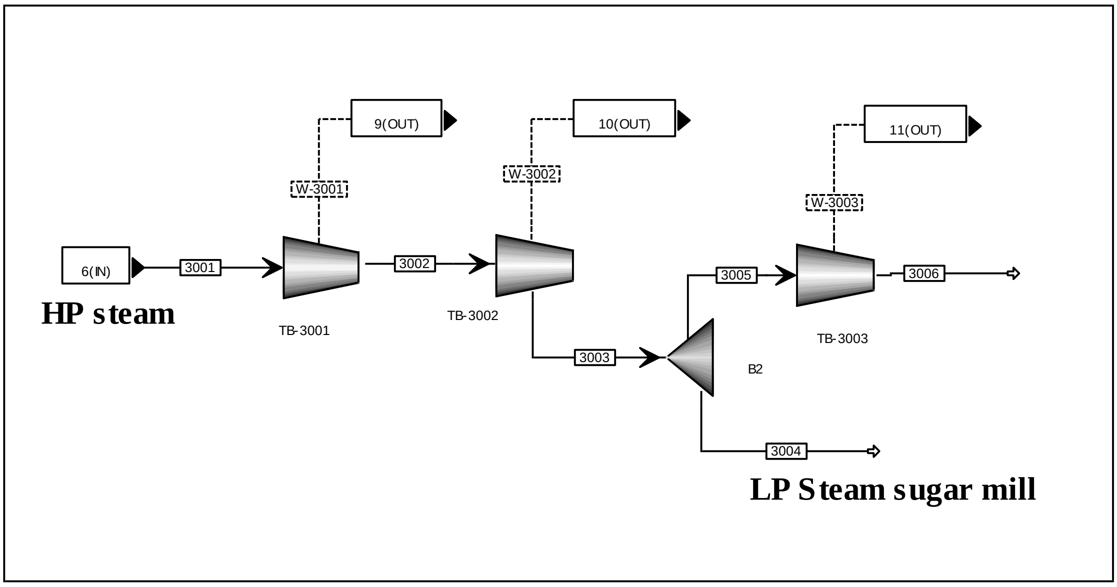
	1001	1002	1003	1004	1005	1006	1007
Mass Flow, tons/hr							
LIGNIN	10.33	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	16.44	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	3.04	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	40.50	0.00	65.26	65.26	65.26	65.26	65.26
ASH	1.46	0.00	1.46	1.46	1.46	1.46	1.46
GALACTAN	0.32	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.08	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	8.10	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.73	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	89.22	32.38	32.38	32.38	32.38	32.38
N_2	0.00	293.82	293.82	293.82	293.82	293.82	293.82
CO_2	0.00	0.00	70.59	70.59	70.59	70.59	70.59
NO_2	0.00	0.00	0.52	0.52	0.52	0.52	0.52
SO_2	0.00	0.00	0.01	0.01	0.01	0.01	0.01
SO_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	81.00	383.04	464.04	464.04	464.04	464.04	464.04
Temperature, ^{o}C	25.00	250.00	1091.77	800.00	500.00	300.90	140.00
Pressure, bar	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Vapour Fraction	0.00	1.00	1.00	1.00	1.00	1.00	1.00
Enthalpy, MJ/hr	-890420.00	79977.62	-810440.00	-980520.00	-1145400.00	-1248100.00	-1327200.00

AREA 2000



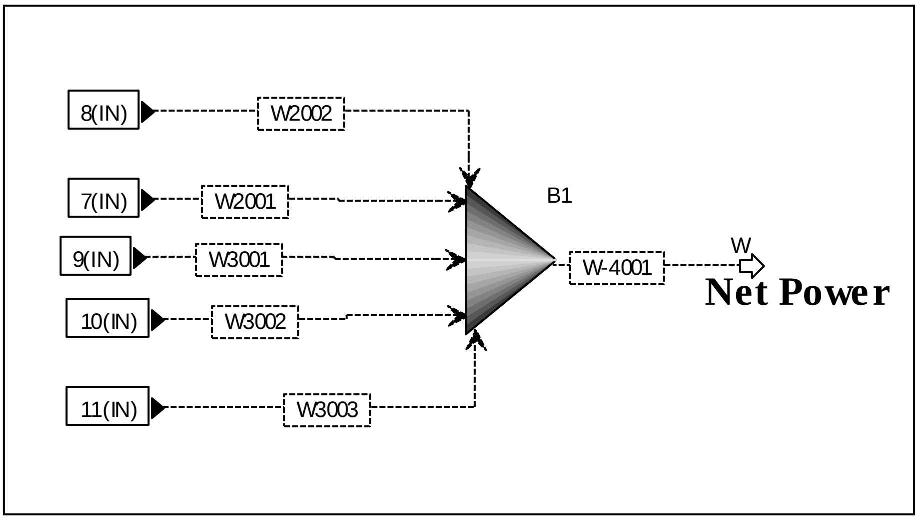
	2001	2002	2003	2004	2005	2006	2007	2008
Mass Flow, tons/hr								
LIGNIN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	0.00	0.00	159.56	159.56	159.56	159.56	159.56
ASH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
GALACTAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	89.22	89.22	89.22	0.00	0.00	0.00	0.00	0.00
N_2	293.82	293.82	293.82	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	383.04	383.04	383.04	159.56	159.56	159.56	159.56	159.56
Temperature, ^oC	25.00	27.39	250.00	80.00	80.84	217.94	250.42	440.00
Pressure, bar	1.00	1.02	1.00	1.30	30.00	40.00	40.00	40.00
Vapour Fraction	1.00	1.00	1.00	0.00	0.00	0.00	0.54	1.00
Enthalpy MJ/hr	0.00	840.84	79977.62	-2263100.00	-2262600.00	-2159900.00	-1995000.00	-1825000.00

AREA 3000



	3001.00	3002.00	3003.00	3004.00	3005.00	3006.00
Mass Flow, tons/hr						
LIGNIN	0.00	0.00	0.00	0.00	0.00	0.00
CELLULOS	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	159.56	159.56	159.56	120.00	39.56	39.56
ASH	0.00	0.00	0.00	0.00	0.00	0.00
GALACTAN	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00
SO_3	0.00	0.00	0.00	0.00	0.00	0.00
S	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow tons/hr	159.56	159.56	159.56	120.00	39.56	39.56
Temperature, ^{o}C	440.00	352.87	129.29	129.29	129.29	60.06
Pressure, bar	40.00	20.00	2.00	2.00	2.00	0.20
Vapour Fraction	1.00	1.00	1.00	1.00	1.00	0.92
Enthalpy, MJ/hr	-1825000.00	-1851000.00	-1914600.00	-1439900.00	-474690.00	-485860.00

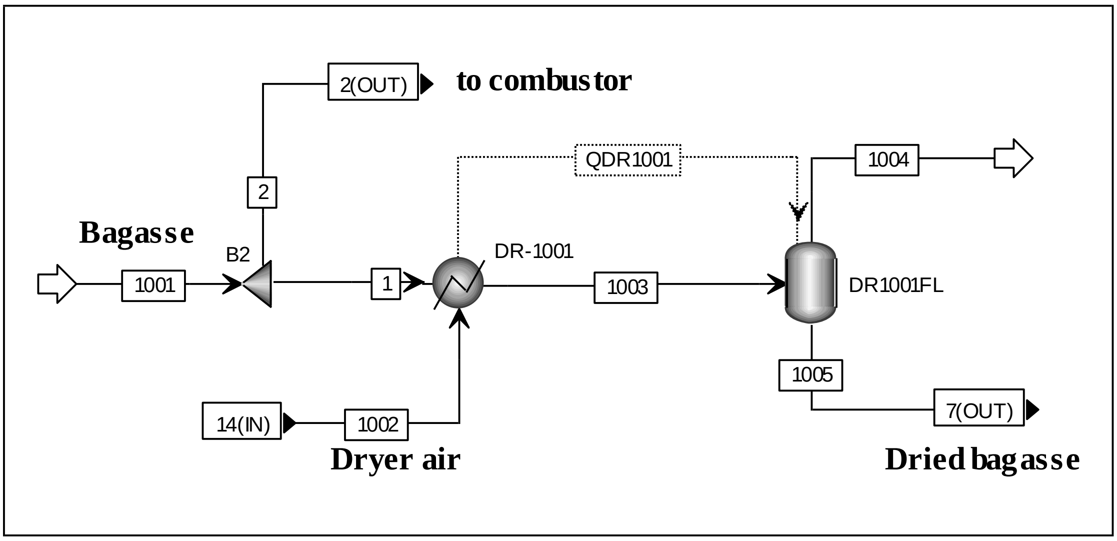
AREA 4000



	W-4001	W2001	W2002	W3001	W3002	W3003
POWER, MW	-27.61	0.14	0.23	-7.22	-17.67	-3.10
SPEED						

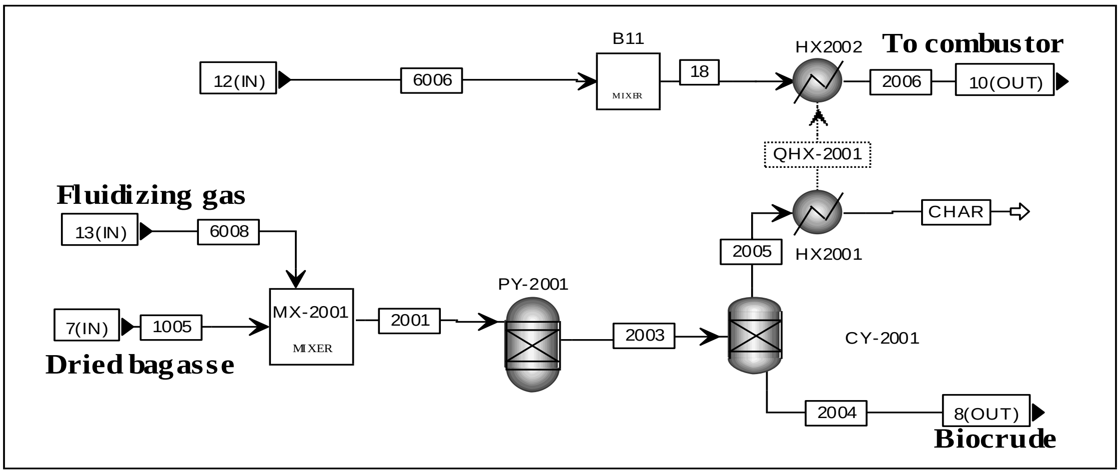
A2 PFD and Mass and Energy Balances for Partial Fast Pyrolysis BMECP (efficient mill)

AREA 1000 - FEED PREPARATION



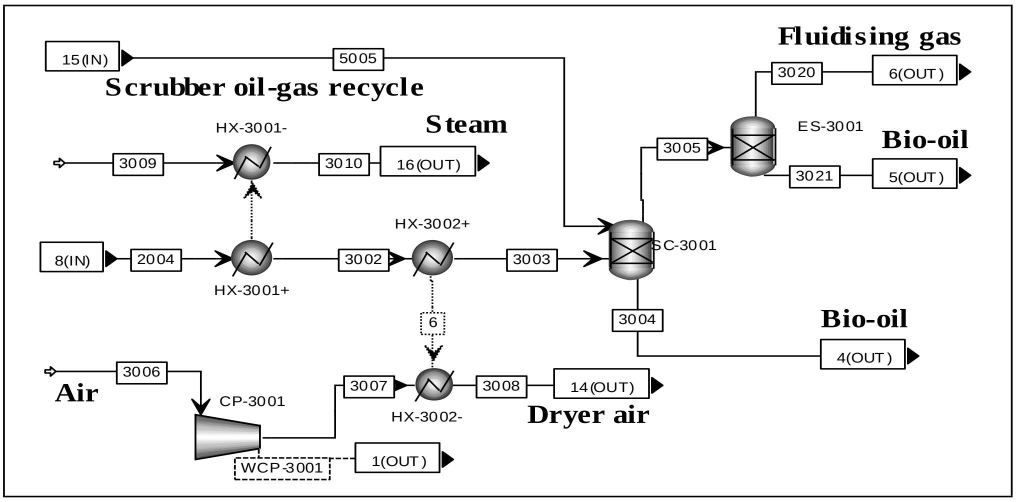
	1	2	1001	1002	1003	1004	1005
Mass Flow, tons/hr							
N_2	0.00	0.00	0.00	100.09	100.09	100.09	0.00
O_2	0.00	0.00	0.00	30.66	30.66	30.66	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	12.15	28.35	40.50	0.88	13.03	11.89	1.14
AR	0.00	0.00	0.00	1.69	1.69	1.69	0.00
ASH	0.44	1.02	1.46	0.00	0.44	0.00	0.44
CELLULOS	4.93	11.51	16.44	0.00	4.93	0.00	4.93
LIGNIN	3.10	7.23	10.33	0.00	3.10	0.00	3.10
XYLAN	2.43	5.67	8.10	0.00	2.43	0.00	2.43
ARABINAN	0.22	0.51	0.73	0.00	0.22	0.00	0.22
MANNAN	0.02	0.06	0.08	0.00	0.02	0.00	0.02
GALACTAN	0.10	0.23	0.32	0.00	0.10	0.00	0.10
EXTRACT	0.91	2.13	3.04	0.00	0.91	0.00	0.91
TETRA-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
DILACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	24.30	56.70	81.00	133.38	157.68	144.38	13.30
Temperature, °C	25.00	25.00	25.00	264.91	110.00	60.00	60.00
Pressure, bar	1.38	1.38	1.38	1.02	1.38	1.38	1.38
Vapour Fraction	0.00	0.00	0.00	1.00	1.00	1.00	0.00
Enthalpy, MJ/hr	-240970.00	-562270.00	-803250.00	18557.19	-212330.00	-140240.00	-82176.02

AREA 2000 - PYROLYSIS



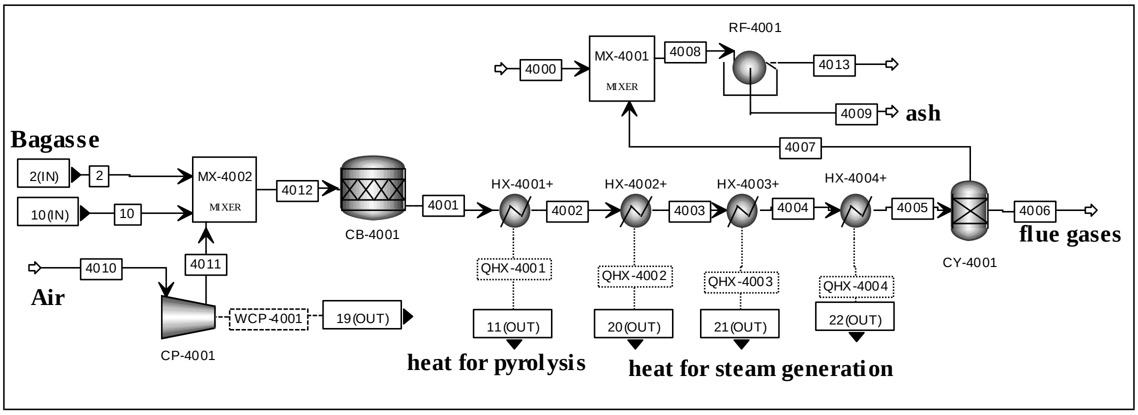
	1005	2001	2003	2004	2005	2006	6006	6008	CHAR
Mass Flow tons/hr									
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	13.12	13.74	13.74	0.00	0.62	0.62	13.12	0.00
CO_2	0.00	18.82	19.71	19.71	0.00	0.89	0.89	18.82	0.00
H_2O	1.14	1.15	2.26	2.26	0.00	0.00	0.00	0.00	0.00
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	1.18	1.24	1.24	0.00	0.06	0.06	1.18	0.00
C_2H_4	0.00	1.18	1.24	1.24	0.00	0.06	0.06	1.18	0.00
ACETACID	0.00	0.00	0.30	0.30	0.00	0.00	0.00	0.00	0.00
ACETOL	0.00	0.00	0.46	0.46	0.00	0.00	0.00	0.00	0.00
GUAIACOL	0.00	0.00	2.17	2.17	0.00	0.00	0.00	0.00	0.00
3:5-X-01	0.00	0.00	0.24	0.24	0.00	0.00	0.00	0.00	0.00
FORMACID	0.00	0.00	0.43	0.43	0.00	0.00	0.00	0.00	0.00
N-PRO-01	0.00	0.00	1.02	1.02	0.00	0.00	0.00	0.00	0.00
PHENOL	0.00	0.00	0.03	0.03	0.00	0.00	0.00	0.00	0.00
TOLUENE	0.00	0.00	0.14	0.14	0.00	0.00	0.00	0.00	0.00
FURFURAL	0.00	0.00	1.19	1.19	0.00	0.00	0.00	0.00	0.00
CHAR	0.00	0.00	1.88	0.00	1.88	0.00	0.00	0.00	1.88
ASH	0.44	0.44	0.44	0.00	0.44	0.00	0.00	0.00	0.44
CELLULOS	4.93	4.93	0.00	0.00	0.00	0.00	0.00	0.00	0.00
LIGNIN	3.10	3.10	0.00	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	2.43	2.43	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.22	0.22	0.00	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.02	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00
GALACTAN	0.10	0.10	0.00	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	0.91	0.91	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TETRA-01	0.00	0.00	0.87	0.87	0.00	0.00	0.00	0.00	0.00
DILACID	0.00	0.00	0.19	0.19	0.00	0.00	0.00	0.00	0.00
Total Flow tons/hr	13.30	47.61	47.61	45.29	2.32	1.62	1.62	34.31	2.32
Temperature, °C	60.00	801.82	500.00	500.00		780.74	91.68	1208.32	
Pressure, bar	1.38	1.38	1.38	1.38	1.38	1.38	3.10	3.10	1.38
Vapour Fraction	0.00	1.00	1.00	1.00		1.00	1.00	1.00	
Enthalpy, MJ/hr	-82176.02	-236040.00	-236040.00	-232810.00	-3224.67	-8180.29	-9449.39	-153860.00	-4493.77

AREA 3000 – QUENCH



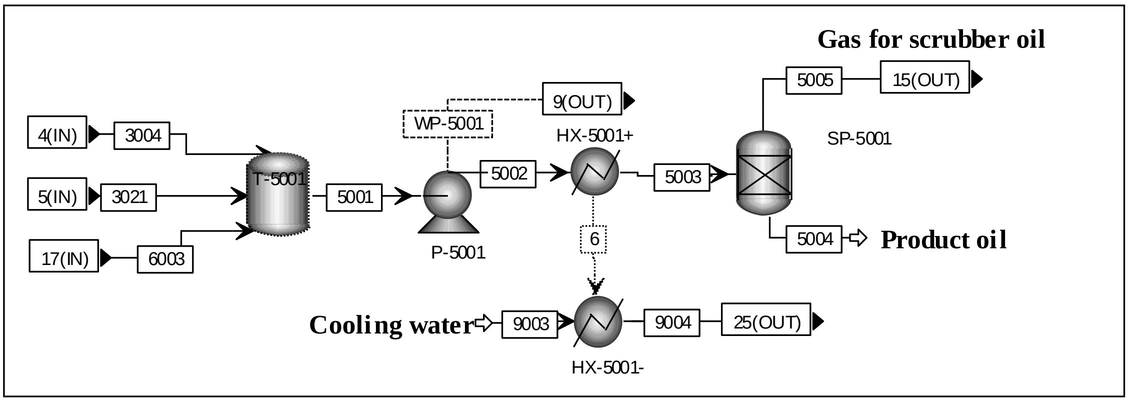
Mass Flow, tons/hr	2004	3002	3003	3004	3005	3006	3007	3008	3009	3010	3020	3021	5005
N_2	0.00	0.00	0.00	0.00	0.00	100.09	100.09	100.09	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	30.66	30.66	30.66	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	13.74	13.74	13.74	0.00	13.74	0.00	0.00	0.00	0.00	0.00	13.74	0.00	0.00
CO_2	19.71	19.71	19.71	0.00	19.71	0.05	0.05	0.05	0.00	0.00	19.71	0.00	0.00
H_2O	2.26	2.26	2.26	0.00	2.66	0.88	0.88	0.88	2.90	2.90	0.00	2.66	0.40
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	1.24	1.24	1.24	0.00	1.24	0.00	0.00	0.00	0.00	0.00	1.24	0.00	0.00
C_2H_4	1.24	1.24	1.24	0.00	1.24	0.00	0.00	0.00	0.00	0.00	1.24	0.00	0.00
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	1.69	1.69	1.69	0.00	0.00	0.00	0.00	0.00
ACETACID	0.30	0.30	0.30	0.28	0.07	0.00	0.00	0.00	0.00	0.00	0.00	0.07	0.05
ACETOL	0.46	0.46	0.46	0.43	0.11	0.00	0.00	0.00	0.00	0.00	0.00	0.11	0.08
GUAIACOL	2.17	2.17	2.17	2.05	0.51	0.00	0.00	0.00	0.00	0.00	0.00	0.51	0.38
3:5-X-01	0.24	0.24	0.24	0.22	0.06	0.00	0.00	0.00	0.00	0.00	0.00	0.06	0.04
FORMACID	0.43	0.43	0.43	0.41	0.10	0.00	0.00	0.00	0.00	0.00	0.00	0.10	0.08
N-PRO-01	1.02	1.02	1.02	0.96	0.24	0.00	0.00	0.00	0.00	0.00	0.00	0.24	0.18
PHENOL	0.03	0.03	0.03	0.03	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.01
TOLUENE	0.14	0.14	0.14	0.13	0.03	0.00	0.00	0.00	0.00	0.00	0.00	0.03	0.03
FURFURAL	1.19	1.19	1.19	1.12	0.28	0.00	0.00	0.00	0.00	0.00	0.00	0.28	0.21
BENZENE	0.05	0.05	0.05	0.05	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.01
TETRA-01	0.87	0.87	0.87	0.87	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
DILACID	0.19	0.19	0.19	0.19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	45.29	45.29	45.29	6.74	40.01	133.38	133.38	133.38	2.90	2.90	35.93	4.08	1.46
Temp., °C	500.00	465.00	28.00	27.02	27.02	25.00	25.80	264.91	80.00	231.62	27.02	27.02	-6.40
Pressure, bar	1.38	1.38	1.38	1.38	1.38	1.01	1.02	1.02	1.38	40.00	1.38	1.38	2.76
Vapour Frac.	1.00	1.00	0.87	0.00	0.89	1.00	1.00	1.00	0.00	0.00	1.00	0.00	0.00
Enthalpy, MJ/hr	-232810.00	-234920.00	-264530.00	-20185.13	-253470.00	-11151.97	-11053.97	18557.19	-41131.75	-39020.42	-212140.00	-42415.99	-8956.47

AREA 4000



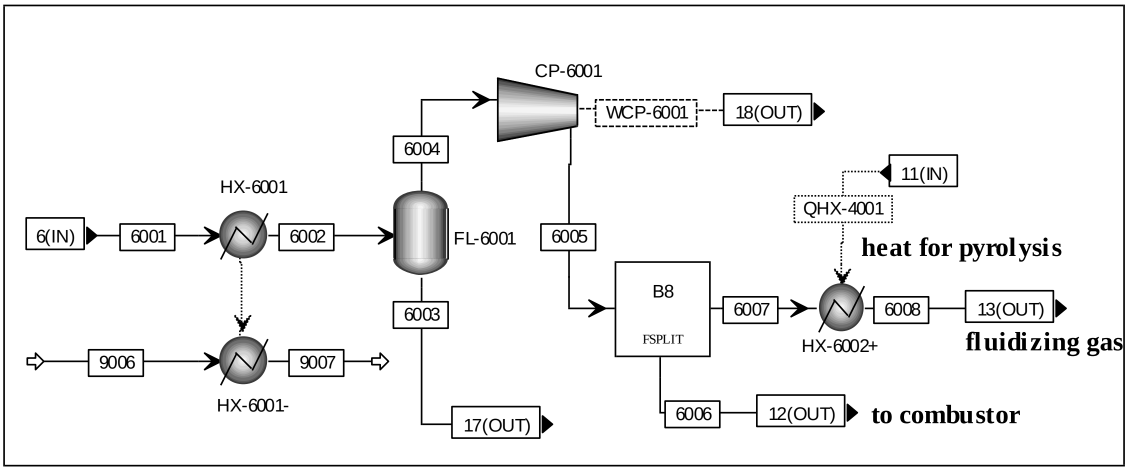
Mass Flow, tons/hr	2	10	4000	4001	4002	4004	4006	4007	4008	4009	4010	4011	4012	4013
N_2	0.00	0.00	0.00	214.94	214.94	214.94	214.94	0.00	0.00	0.00	214.94	214.94	214.94	0.00
O_2	0.00	0.00	0.00	24.52	24.52	24.52	24.52	0.00	0.00	0.00	65.86	65.86	65.86	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	0.62	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.62	0.00
$N0_2$	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.89	0.00	50.79	50.79	50.79	50.79	0.00	0.00	0.00	0.14	0.14	1.03	0.00
H_2O	28.35	0.00	1.23	47.75	47.75	47.75	47.75	0.00	1.23	0.00	1.89	1.89	30.24	1.23
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.06	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.06	0.00
C_2H_4	0.00	0.06	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.06	0.00
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	3.64	3.64	3.64	3.64	0.00	0.00	0.00	3.64	3.64	3.64	0.00
ACETACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
GUAIACOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
3:5-X-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FORMACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N-PRO-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PHENOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOLUENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FURFURAL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
BENZENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CHAR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ASH	1.02	0.00	0.00	1.02	1.02	1.02	0.00	1.02	1.02	1.02	0.00	0.00	1.02	0.00
CELLULOS	11.51	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	11.51	0.00
LIGNIN	7.23	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	7.23	0.00
XYLAN	5.67	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	5.67	0.00
ARABINAN	0.51	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.51	0.00
MANNAN	0.06	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.06	0.00
GALACTAN	0.23	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.23	0.00
EXTRACT	2.13	0.00	0.00	2.13	2.13	2.13	2.13	0.00	0.00	0.00	0.00	0.00	2.13	0.00
TETRA-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
DILACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow tons/hr	56.70	1.62	1.23	344.79	344.79	344.79	343.77	1.02	2.25	1.02	286.48	286.48	344.79	1.23
Temp., ^oC	25.00	780.74	25.00	1095.02	987.62	500.00	120.00		40.00		150.00	151.12	85.46	40.00
Pressure, bar	1.38	1.38	1.38	1.02	1.02	1.02	1.02	1.02	1.02		1.01	1.02	14.80	1.02
Vapour Frac.	0.00	1.00	0.00	1.00	1.00	1.00	1.00		0.00		1.00	1.00	0.89	0.00
Enthalpy, MJ/hr	-562270.00	-8180.29	-17730.25	-561360.00	-608030.00	-808670.00	-939130.00	-10407.72	-28137.97	-10473.25	8797.81	9096.54	-561360.00	-17664.71

AREA 5000 – OIL RECOVERY



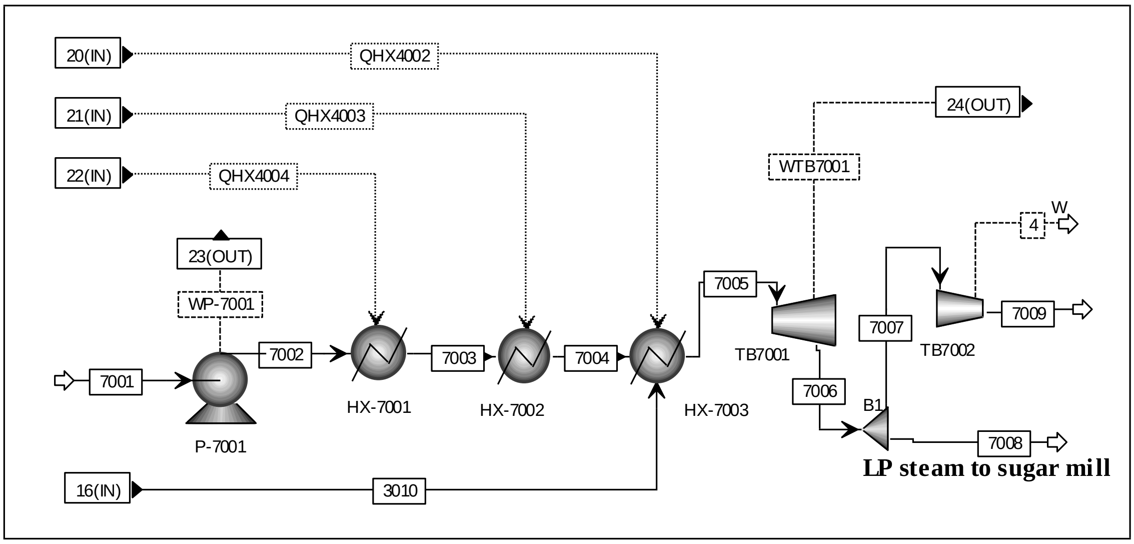
Mass Flow, tons/hr	3004	3021	5001	5002	5003	5004	5005	6003	9003	9004
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
$N0_2$	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	2.66	2.66	2.66	2.66	2.26	0.40	0.00	16.32	16.32
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_2H_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETACID	0.28	0.07	0.36	0.36	0.36	0.30	0.05	0.00	0.00	0.00
ACETOL	0.43	0.11	0.54	0.54	0.54	0.46	0.08	0.00	0.00	0.00
GUAIACOL	2.05	0.51	2.56	2.56	2.56	2.17	0.38	0.00	0.00	0.00
3:5-X-01	0.22	0.06	0.28	0.28	0.28	0.24	0.04	0.00	0.00	0.00
FORMACID	0.41	0.10	0.51	0.51	0.51	0.43	0.08	0.00	0.00	0.00
N-PRO-01	0.96	0.24	1.20	1.20	1.20	1.02	0.18	0.00	0.00	0.00
PHENOL	0.03	0.01	0.03	0.03	0.03	0.03	0.01	0.00	0.00	0.00
TOLUENE	0.13	0.03	0.17	0.17	0.17	0.14	0.03	0.00	0.00	0.00
FURFURAL	1.12	0.28	1.40	1.40	1.40	1.19	0.21	0.00	0.00	0.00
BENZENE	0.05	0.01	0.06	0.06	0.06	0.05	0.01	0.00	0.00	0.00
TETRA-01	0.87	0.00	0.87	0.87	0.87	0.87	0.00	0.00	0.00	0.00
DILACID	0.19	0.00	0.19	0.19	0.19	0.19	0.00	0.00	0.00	0.00
Total Flow, tons/hr	6.74	4.08	10.82	10.82	10.82	9.36	1.46	0.00	16.32	16.32
Temperature, ^{o}C	27.02	27.02	26.15	26.25	25.00	-6.40	-6.40		21.10	21.59
Pressure, bar	1.38	1.38	1.79	2.76	2.76	2.76	2.76	1.38	1.03	1.03
Vapour Frac	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00
Enthalpy, MJ/hr	-20185.13	-42415.99	-62601.13	-62598.96	-62626.81	-54354.30	-8956.43	0.00	-234990.00	-234970.00

AREA 6000 – RECYCLE



	6001	6002	6003	6004	6005	6006	6007	6008	9006	9007
Mass Flow, tons/hr										
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	13.74	13.74	0.00	13.74	13.74	0.62	13.12	13.12	0.00	0.00
$N0_2$										
CO_2	19.71	19.71	0.00	19.71	19.71	0.89	18.82	18.82	0.00	0.00
H_2O	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	70.77	70.77
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	1.24	1.24	0.00	1.24	1.24	0.06	1.18	1.18	0.00	0.00
C_2H_4	1.24	1.24	0.00	1.24	1.24	0.06	1.18	1.18	0.00	0.00
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
GUAIACOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
3:5-X-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FORMACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N-PRO-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PHENOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOLUENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FURFURAL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
BENZENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TETRA-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
DILACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	35.93	35.93	0.00	35.93	35.93	1.62	34.31	34.31	70.77	70.77
Temperature, ^{o}C	27.02	7.00		7.00	91.68	91.68	91.68	1208.32	4.00	6.68
Pressure, bar	1.38	1.38	1.38	1.38	3.10	3.10	3.10	3.10	1.03	1.03
Vapour Frac	1.00	1.00		1.00	1.00	1.00	1.00	1.00	0.00	0.00
Enthalpy, MJ/hr	-212140.00	-212780.00	0.00	-212780.00	-209990.00	-9449.39	-200540.00	-153860.00	-1023100.00	-1022500.00

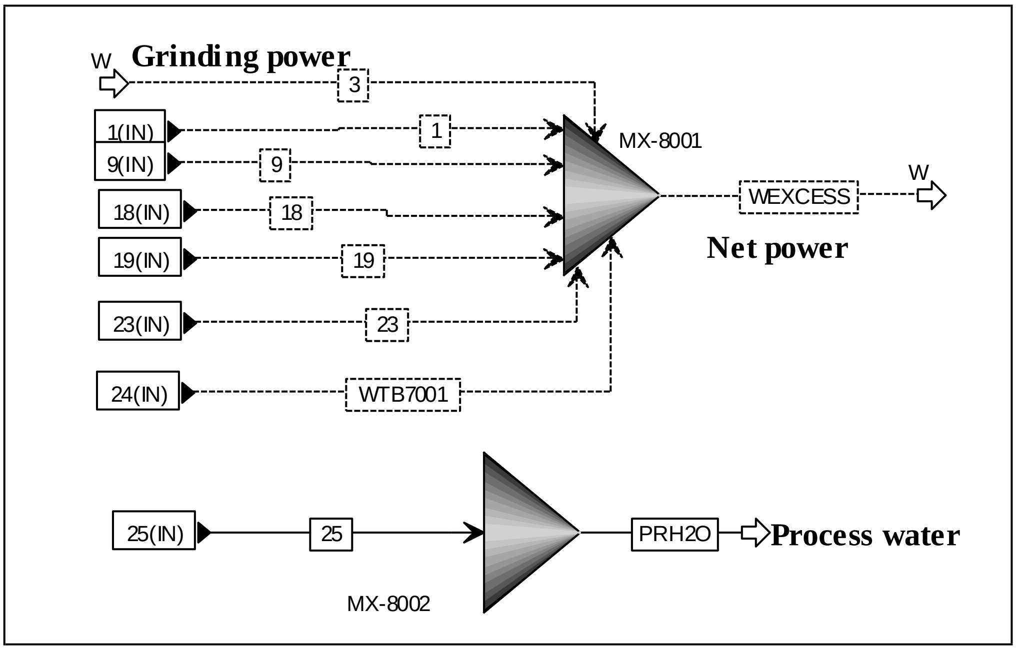
AREA 7000 – STEAM AND POWER CYCLE



LP steam to sugar mill

	3010	7001	7002	7003	7004	7005	7011
Mass Flow, tons/hr							
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00
$N0_2$	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	2.90	125.82	125.82	125.82	125.82	128.72	128.72
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_2H_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N-PRO-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PHENOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOLUENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FURFURAL	0.00	0.00	0.00	0.00	0.00	0.00	0.00
BENZENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TETRA-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
DILACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	2.90	125.82	125.82	125.82	125.82	128.72	128.72
Temperature, °C	231.62	80.00	80.31	250.42	250.42	400.00	120.27
Pressure, bar	40.00	1.38	10.41	40.00	40.00	40.00	2.00
Vapour Frac	0.00	0.00	0.00	0.18	0.59	1.00	0.99
Enthalpy, MJ/hr	-39020.42	-1784600.00	-1784400.00	-1643600.00	-1563800.00	-1481900.00	-1549500.00

AREA 8000 – UTILITIES

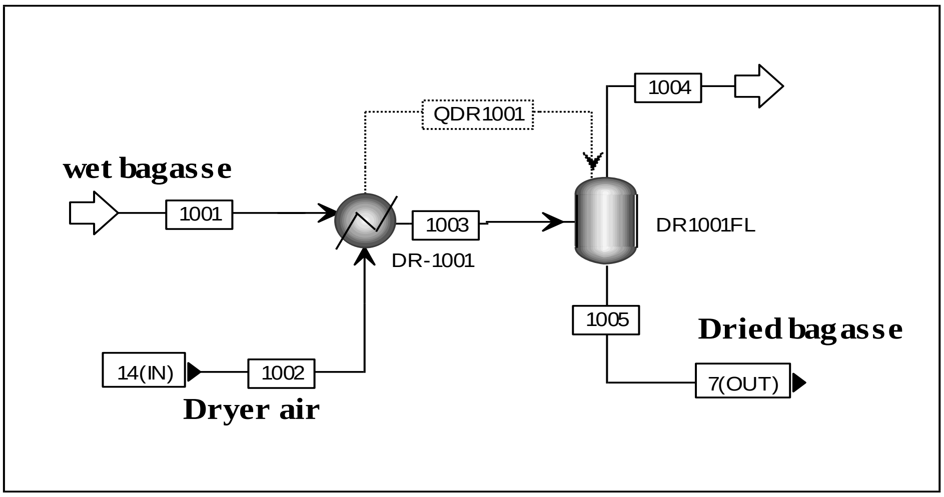


	1	3	9	18	19	23	WEXCESS	WTB7001
POWER, MW	0.03	1.26	0.00	0.78	0.08	0.04	-16.58	-18.77
SPEED								

	25	PRH2O
Mass Flow, tons/hr		
H_2O	16.32	16.32
Total Flow tons/hr	16.32	16.32
Temperature, ^{o}C	21.59	21.59
Pressure, bar	1.03	1.03
Vapour Frac	0.00	0.00
Enthalpy, MJ/hr	-234970.00	-234970.00

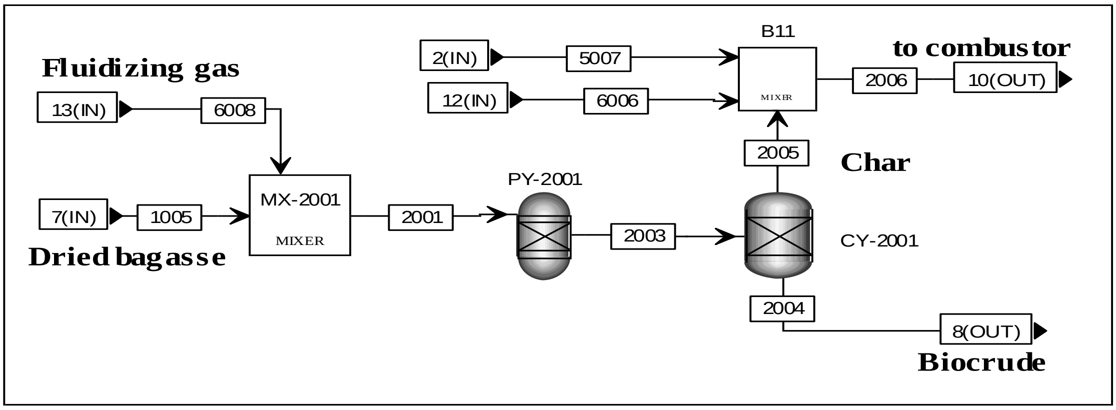
A.3 PFD and Mass and Energy Balances of Pure Fast Pyrolysis BMECP (efficient mill)

AREA 1000 – FEED PREPARATION



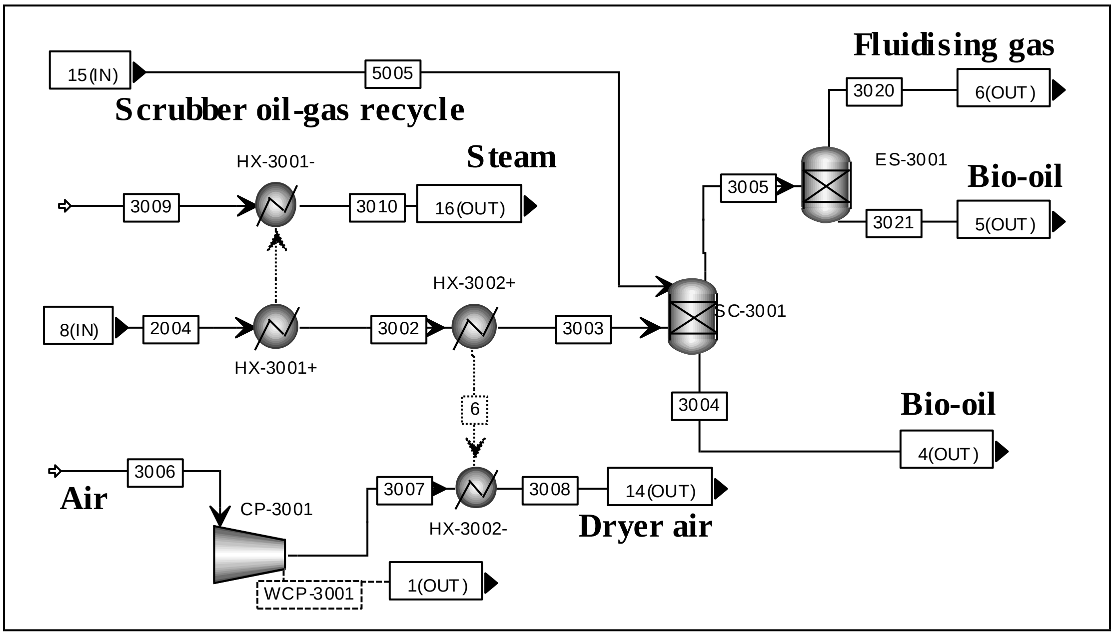
	1001	1002	1003	1004	1005
Mass Flow, tons/hr					
N_2	0.00	333.55	333.55	333.54	0.01
O_2	0.00	102.19	102.19	102.19	0.00
H_2	0.00	0.00	0.00	0.00	0.00
CO	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.18	0.18	0.18	0.00
H_2O	40.50	2.93	43.43	39.62	3.81
AR	0.00	5.65	5.65	5.64	0.00
CHAR	0.00	0.00	0.00	0.00	0.00
ASH	1.46	0.00	1.46	0.00	1.46
CELLULOS	16.44	0.00	16.44	0.00	16.44
LIGNIN	10.33	0.00	10.33	0.00	10.33
XYLAN	8.10	0.00	8.10	0.00	8.10
ARABINAN	0.73	0.00	0.73	0.00	0.73
MANNAN	0.08	0.00	0.08	0.00	0.08
GALACTAN	0.32	0.00	0.32	0.00	0.32
EXTRACT	3.04	0.00	3.04	0.00	3.04
TETRA-01	0.00	0.00	0.00	0.00	0.00
DILACID	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	81.00	444.49	525.49	481.17	44.32
Temperature, °C	25.00	264.94	110.00	60.00	60.00
Pressure, bar	1.38	1.02	1.38	1.38	1.38
Vapour Frac	0.00	1.00	1.00	1.00	0.00
Enthalpy, MJ/hr	-803250.00	61855.39	-707780.00	-467430.00	-273970.00

AREA 2000- PYROLYSIS



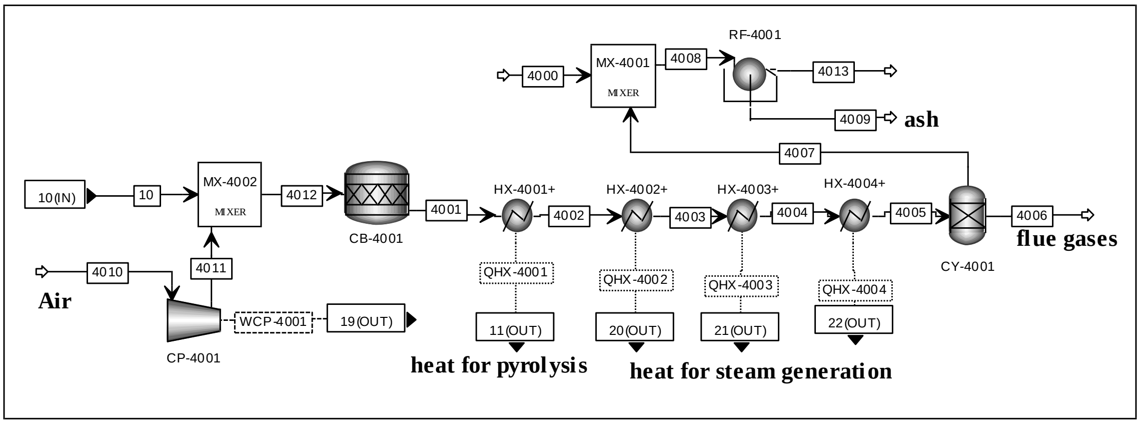
	1005	2001	2003	2004	2005	2006	5007	6006	6008
Mass Flow, tons/hr									
N_2	0.01	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	43.73	45.79	45.79	0.00	2.06	0.00	2.06	43.73
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	62.74	65.70	65.70	0.00	2.96	0.00	2.96	62.74
H_2O	3.81	3.82	7.55	7.55	0.00	3.51	3.51	0.00	0.01
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	3.94	4.13	4.13	0.00	0.19	0.00	0.19	3.94
C_2H_4	0.00	3.94	4.13	4.13	0.00	0.19	0.00	0.19	3.94
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETACID	0.00	0.00	1.01	1.01	0.00	0.47	0.47	0.00	0.00
ACETOL	0.00	0.00	1.53	1.53	0.00	0.71	0.71	0.00	0.00
GUAIACOL	0.00	0.00	7.24	7.24	0.00	3.37	3.37	0.00	0.00
3:5-X-01	0.00	0.00	0.79	0.79	0.00	0.37	0.37	0.00	0.00
FORMACID	0.00	0.00	1.45	1.45	0.00	0.67	0.67	0.00	0.00
N-PRO-01	0.00	0.00	3.41	3.41	0.00	1.59	1.59	0.00	0.00
PHENOL	0.00	0.00	0.10	0.10	0.00	0.05	0.05	0.00	0.00
TOLUENE	0.00	0.00	0.47	0.47	0.00	0.22	0.22	0.00	0.00
FURFURAL	0.00	0.00	3.96	3.96	0.00	1.84	1.84	0.00	0.00
BENZENE	0.00	0.00	0.16	0.16	0.00	0.07	0.07	0.00	0.00
CHAR	0.00	0.00	6.27	0.00	6.27	6.27	0.00	0.00	0.00
ASH	1.46	1.46	1.46	0.00	1.46	1.46	0.00	0.00	0.00
CELLULOS	16.44	16.44	0.00	0.00	0.00	0.00	0.00	0.00	0.00
LIGNIN	10.33	10.33	0.00	0.00	0.00	0.00	0.00	0.00	0.00
XYLAN	8.10	8.10	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ARABINAN	0.73	0.73	0.00	0.00	0.00	0.00	0.00	0.00	0.00
MANNAN	0.08	0.08	0.00	0.00	0.00	0.00	0.00	0.00	0.00
GALACTAN	0.32	0.32	0.00	0.00	0.00	0.00	0.00	0.00	0.00
EXTRACT	3.04	3.04	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TETRA-01	0.00	0.00	2.90	2.90	0.00	1.35	1.35	0.00	0.00
DILACID	0.00	0.00	0.64	0.64	0.00	0.30	0.30	0.00	0.00
Total Flow, tons/hr	44.32	158.68	158.68	150.95	7.73	27.63	14.51	5.39	114.36
Temperature, °C	60.00	801.77	499.90	499.90		76.14	25.00	91.68	1208.31
Pressure, bar	1.38	1.38	1.38	1.38	1.38	1.38	2.76	3.10	3.10
Vapour Frac	0.00	1.00	1.00	1.00		0.49	0.00	1.00	1.00
Enthalpy, MJ/hr	-273970.00	-786780.00	-786780.00	-776030.00	-10750.25	-125580.00	-83333.59	-31496.98	-512810.00

AREA 3000 – QUENCH



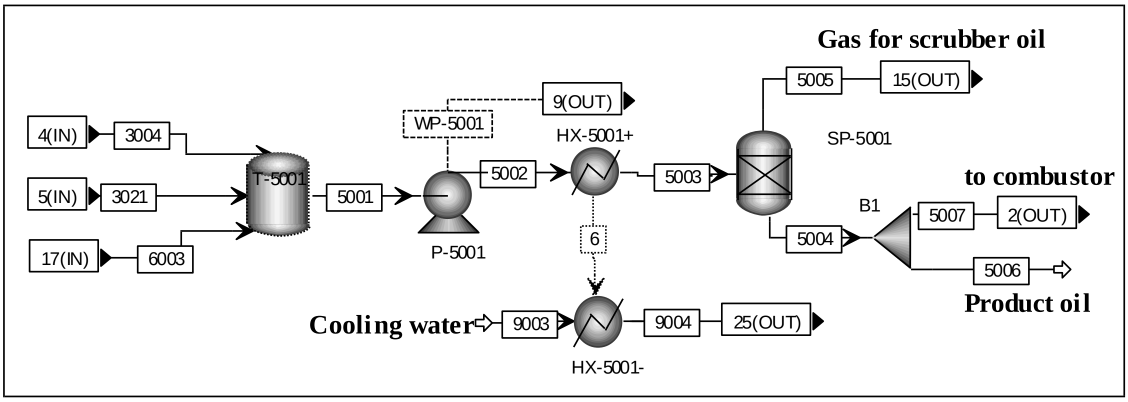
Mass Flow, tons/hr	2004	3002	3003	3004	3005	3006	3007	3008	3009	3010	3020	3021	5005
N_2	0.00	0.00	0.00	0.00	0.00	333.55	333.55	333.55	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	102.19	102.19	102.19	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	45.79	45.79	45.79	0.00	45.79	0.00	0.00	0.00	0.00	0.00	45.79	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	65.70	65.70	65.70	0.00	65.70	0.18	0.18	0.18	0.00	0.00	65.70	0.00	0.00
H_2O	7.55	7.55	7.55	0.00	8.88	2.93	2.93	2.93	2.90	2.90	0.01	8.87	1.33
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	4.13	4.13	4.13	0.00	4.13	0.00	0.00	0.00	0.00	0.00	4.13	0.00	0.00
C_2H_4	4.13	4.13	4.13	0.00	4.13	0.00	0.00	0.00	0.00	0.00	4.13	0.00	0.00
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	5.65	5.65	5.65	0.00	0.00	0.00	0.00	0.00
ACETACID	1.01	1.01	1.01	0.95	0.24	0.00	0.00	0.00	0.00	0.00	0.00	0.24	0.18
ACETOL	1.53	1.53	1.53	1.44	0.36	0.00	0.00	0.00	0.00	0.00	0.00	0.36	0.27
GUAIACOL	7.24	7.24	7.24	6.82	1.70	0.00	0.00	0.00	0.00	0.00	0.00	1.70	1.28
3:5-X-01	0.79	0.79	0.79	0.75	0.19	0.00	0.00	0.00	0.00	0.00	0.00	0.19	0.14
FORMACID	1.45	1.45	1.45	1.36	0.34	0.00	0.00	0.00	0.00	0.00	0.00	0.34	0.26
N-PRO-01	3.41	3.41	3.41	3.21	0.80	0.00	0.00	0.00	0.00	0.00	0.00	0.80	0.60
PHENOL	0.10	0.10	0.10	0.09	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.02
TOLUENE	0.47	0.47	0.47	0.45	0.11	0.00	0.00	0.00	0.00	0.00	0.00	0.11	0.08
FURFURAL	3.96	3.96	3.96	3.73	0.93	0.00	0.00	0.00	0.00	0.00	0.00	0.93	0.70
BENZENE	0.16	0.16	0.16	0.15	0.04	0.00	0.00	0.00	0.00	0.00	0.00	0.04	0.03
TETRA-01	2.90	2.90	2.90	2.90	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
DILACID	0.64	0.64	0.64	0.64	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	150.95	150.95	150.95	22.48	133.35	444.49	444.49	444.49	2.90	2.90	119.75	13.60	4.88
Temp., °C	499.90	465.00	28.00	27.87	27.87	25.00	25.80	264.94	80.00	263.17	27.87	27.87	25.00
Pressure, bar	1.38	1.38	1.38	1.38	1.38	1.01	1.02	1.02	1.38	40.00	1.38	1.38	2.76
Vapour Frac	1.00	1.00	0.87	0.00	0.89	1.00	1.00	1.00	0.00	1.00	1.00	0.00	0.00
Enthalpy, MJ/hr	-776030.00	-783050.00	-881750.00	-67254.80	-844580.00	-37164.97	-36838.37	61855.39	-41131.75	-34114.19	-707010.00	-141350.00	-29536.04

AREA 4000 – HEAT RECOVERY



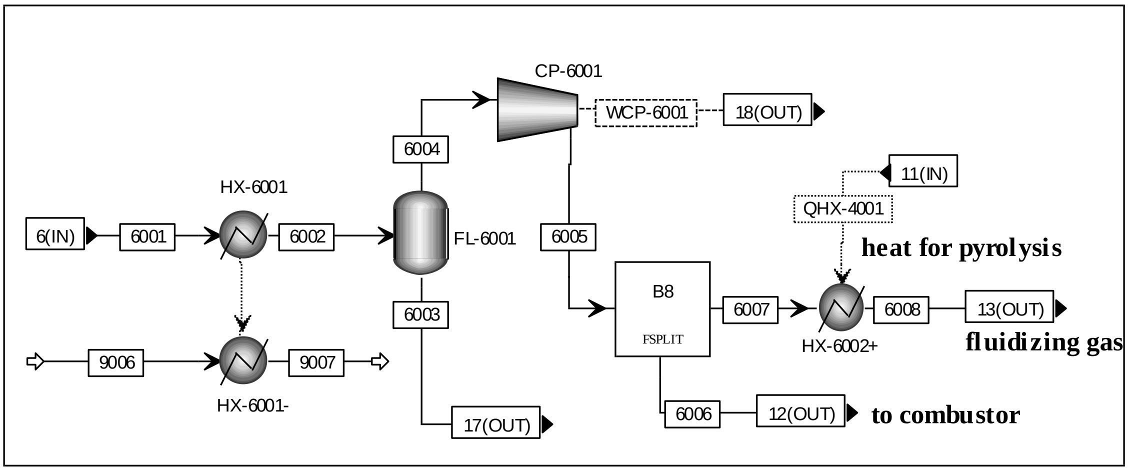
	10	4000	4001	4002	4004	4006	4007	4008	4009	4010	4011	4012	4013
Mass Flow, tons/hr													
N_2	0.00	0.00	200.96	200.96	200.96	200.96	0.00	0.00	0.00	200.96	200.96	200.96	0.00
O_2	0.00	0.00	21.15	21.15	21.15	21.15	0.00	0.00	0.00	61.58	61.58	61.58	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	2.06	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	2.06	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	2.96	0.00	55.78	55.78	55.78	55.78	0.00	0.00	0.00	0.13	0.13	3.09	0.00
H_2O	3.51	1.76	12.72	12.72	12.72	12.72	0.00	1.76	0.00	1.77	1.77	5.27	1.76
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.19	0.00
C_2H_4	0.19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.19	0.00
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	3.40	3.40	3.40	3.40	0.00	0.00	0.00	3.40	3.40	3.40	0.00
ACETACID	0.47	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.47	0.00
ACETOL	0.71	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.71	0.00
GUAIACOL	3.37	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3.37	0.00
3:5-X-01	0.37	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.37	0.00
FORMACID	0.67	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.67	0.00
N-PRO-01	1.59	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.59	0.00
PHENOL	0.05	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.05	0.00
TOLUENE	0.22	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.22	0.00
FURFURAL	1.84	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.84	0.00
BENZENE	0.07	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.07	0.00
CHAR	473.82	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	473.82	0.00
ASH	23.59	0.00	23.59	23.59	23.59	0.00	23.59	23.59	23.59	0.00	0.00	23.59	0.00
TETRA-01	1.35	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.35	0.00
DILACID	0.30	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.30	0.00
Total Flow, tons/hr	27.63	1.76	295.47	295.47	295.47	294.01	1.46	3.22	1.46	267.84	267.84	295.47	1.76
Temperature, ^oC	76.14	25.00	1581.70	1141.00	500.00	120.00		40.00		150.00	151.12	115.77	40.00
Pressure, bar	1.38	1.38	1.02	1.02	1.02	1.02	1.02	1.02		1.01	1.02	14.80	1.02
Vapour Frac	0.49	0.00	1.00	1.00	1.00	1.00		0.00		1.00	1.00	1.00	0.00
Enthalpy, MJ/hr	-125580.00	-25331.89	-117080.00	-272660.00	-484060.00	-581070.00	-14868.17	-40200.05	-14961.79	8225.48	8504.77	-117080.00	-25238.27

AREA 5000 – OIL RECOVERY



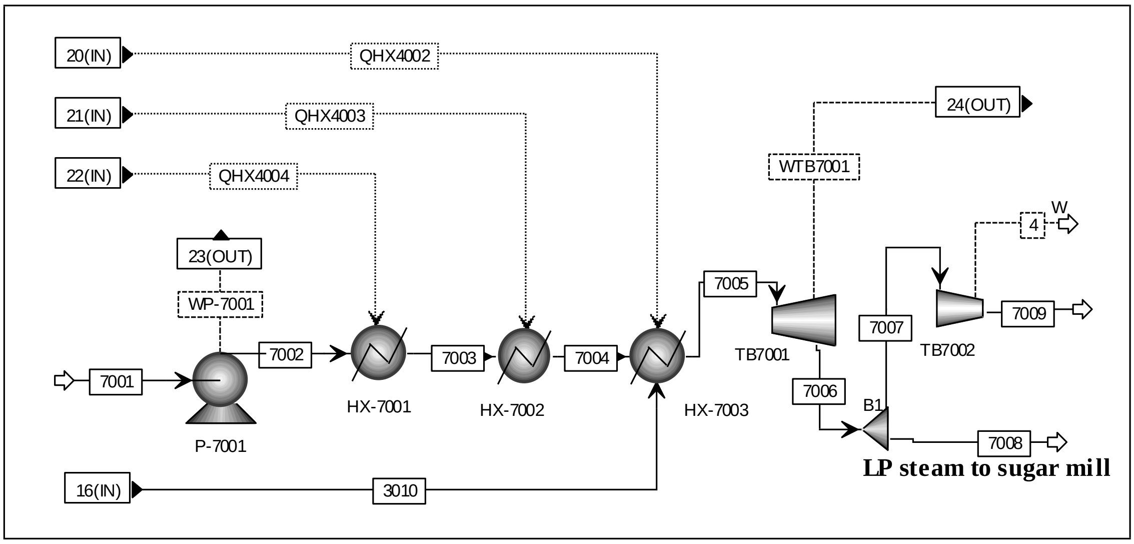
	3004	3021	5001	5002	5003	5004	5005	5006	5007	6003	9003	9004
Mass Flow, tons/hr												
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	8.87	8.87	8.87	8.87	7.54	1.33	4.03	3.51	0.00	16.32	16.32
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_2H_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETACID	0.95	0.24	1.18	1.18	1.18	1.01	0.18	0.54	0.47	0.00	0.00	0.00
ACETOL	1.44	0.36	1.79	1.79	1.79	1.52	0.27	0.82	0.71	0.00	0.00	0.00
GUAIACOL	6.82	1.70	8.52	8.52	8.52	7.24	1.28	3.88	3.37	0.00	0.00	0.00
3:5-X-01	0.75	0.19	0.93	0.93	0.93	0.79	0.14	0.42	0.37	0.00	0.00	0.00
FORMACID	1.36	0.34	1.70	1.70	1.70	1.45	0.26	0.77	0.67	0.00	0.00	0.00
N-PRO-01	3.21	0.80	4.01	4.01	4.01	3.41	0.60	1.82	1.59	0.00	0.00	0.00
PHENOL	0.09	0.02	0.11	0.11	0.11	0.10	0.02	0.05	0.05	0.00	0.00	0.00
TOLUENE	0.45	0.11	0.56	0.56	0.56	0.47	0.08	0.25	0.22	0.00	0.00	0.00
FURFURAL	3.73	0.93	4.66	4.66	4.66	3.96	0.70	2.12	1.84	0.00	0.00	0.00
BENZENE	0.15	0.04	0.19	0.19	0.19	0.16	0.03	0.09	0.07	0.00	0.00	0.00
TETRA-01	2.90	0.00	2.90	2.90	2.90	2.90	0.00	1.55	1.35	0.00	0.00	0.00
DILACID	0.64	0.00	0.64	0.64	0.64	0.64	0.00	0.34	0.30	0.00	0.00	0.00
Total Flow, tons/hr	22.48	13.60	36.08	36.08	36.08	31.20	4.88	16.69	14.51	0.00	16.32	16.32
Temperature, °C	27.87	27.87	26.96	27.03	25.00	25.00	25.00	25.00	25.00		21.10	23.75
Pressure, bar	1.38	1.38	1.79	2.76	2.76	2.76	2.76	2.76	2.76	1.38	1.03	1.03
Vapour Frac	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00
Enthalpy, MJ/hr	-67254.80	-141350.00	-208600.00	-208600.00	-208750.00	-179210.00	-29536.03	-95878.43	-83333.59	0.00	-234990.00	-234840.00

AREA 6000 – RECYCLE



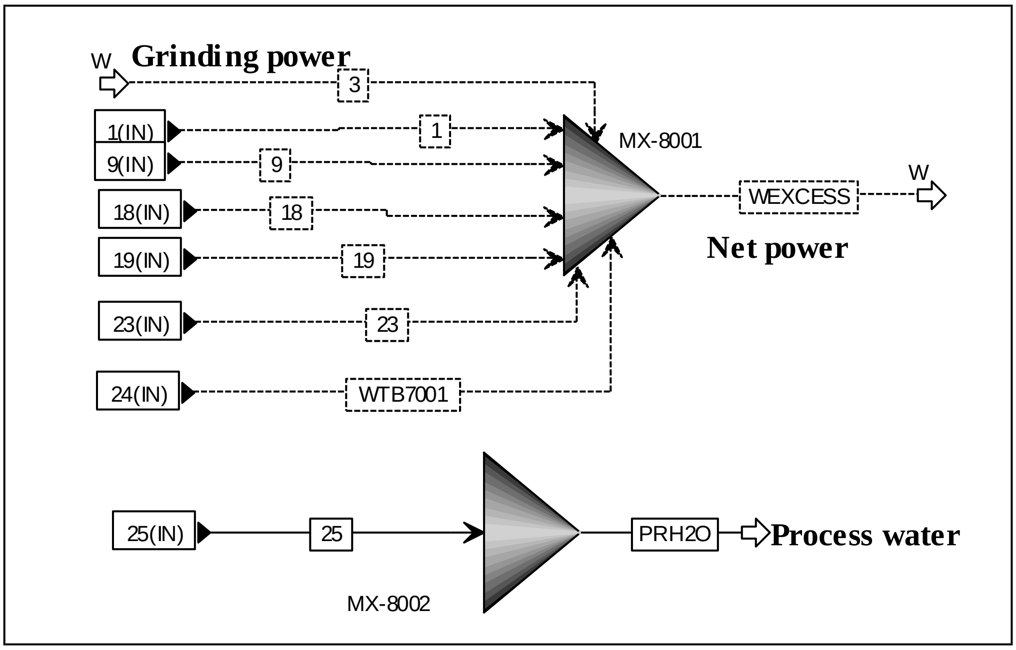
	6001	6002	6003	6004	6005	6006	6007	6008	9006	9007
Mass Flow, tons/hr										
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	45.79	45.79	0.00	45.79	45.79	2.06	43.73	43.73	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	65.70	65.70	0.00	65.70	65.70	2.96	62.74	62.74	0.00	0.00
H_2O	0.01	0.01	0.00	0.01	0.01	0.00	0.01	0.01	70.77	70.77
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	4.13	4.13	0.00	4.13	4.13	0.19	3.94	3.94	0.00	0.00
C_2H_4	4.13	4.13	0.00	4.13	4.13	0.19	3.94	3.94	0.00	0.00
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
GUAIACOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
3:5-X-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FORMACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N-PRO-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PHENOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOLUENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FURFURAL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
BENZENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TETRA-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
DILACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	119.75	119.75	0.00	119.75	119.75	5.39	114.37	114.37	70.77	70.77
Temperature, °C	27.87	7.00		7.00	91.68	91.68	91.68	1208.31	4.00	13.27
Pressure, bar	1.38	1.38	1.38	1.38	3.10	3.10	3.10	3.10	1.03	1.03
Vapour Frac	1.00	1.00		1.00	1.00	1.00	1.00	1.00	0.00	0.00
Enthalpy, MJ/hr	-707010.00	-709240.00	0.00	-709240.00	-699930.00	-31496.98	-668440.00	-512850.00	-1023100.00	-1020900.00

AREA 7000 – STEAM AND POWER CYCLE



	3010	7001	7002	7003	7004	7005	7011
Mass Flow, tons/hr							
N_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	2.90	120.83	120.83	120.83	120.83	123.73	123.73
H_3N	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_2H_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_6-2	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00
ACETOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00
GUAIACOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00
3:5-X-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FORMACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N-PRO-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PHENOL	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOLUENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FURFURAL	0.00	0.00	0.00	0.00	0.00	0.00	0.00
BENZENE	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CHAR	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TETRA-01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
DILACID	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow, tons/hr	2.90	120.83	120.83	120.83	120.83	123.73	123.73
Temperature, ^oC	263.17	80.00	80.31	250.42	250.42	400.00	120.27
Pressure, bar	40.00	1.38	10.41	40.00	40.00	40.00	2.00
Vapour Frac	1.00	0.00	0.00	0.05	0.39	1.00	0.99
Enthalpy, MJ/hr	-34114.19	-1713800.00	-1713700.00	-1601800.00	-1538500.00	-1424500.00	-1489500.00

AREA 8000 - UTILITIES



	1	3	9	18	19	23	WEXCESS	WTB7001
POWER MW	0.09	4.19	0.00	2.59	0.08	0.04	-11.05	-18.04
SPEED								

	25	PRH2O
Mass Flow, tons/hr		
H_2O	16.32	16.32
Total Flow, tons/hr	16.32	16.32
Temperature, ^oC	23.75	23.75
Pressure, bar	1.03	1.03
Vapour Fraction	0.00	0.00
Enthalpy, MJ/hr	-234840.00	-234840.00

APPENDIX B – BIOMASS COMPOSTION AND PROPERTY DATA

Table B.1: Chemical Formulas and Property Data Sources for Biomass Components Used in AspenPlus® Process Models

Component name	Chemical Formula	Properties Source
Cellulose	$C_5H_{10}O_5$	NREL in-house databank
Galactan	$C_5H_{10}O_5$	Cellulose
Xylan	$C_5H_8O_4$	NREL in-house databank
Arabinan	$C_5H_8O_4$	Xylan
Mannan	$C_5H_8O_4$	Xylan
Lignin	$C_{10}H_{13.9}O_{1.3}$	NREL in-house databank
Extractives		NREL in-house databank

Table B.2: Biomass Composition Used in Aspen Plus® Process Models

Component	Weight % (dry and extractive free basis)
Cellulose	43.89
Galactan	0.86
Mannan	0.22
Xylan	21.62
Arabinan	1.95
Lignin	27.57
Ash	3.89
Extractives	7.50 (dry basis only)
LHV @ 50% moisture [a]	7.46 MJ/kg

[a] Calculated based on the formulae of Phyllis (2005) as discussed in the text.

APPENDIX C – RELATIONSHIP BETWEEN TEMPERATURE AND EFFICIENCY

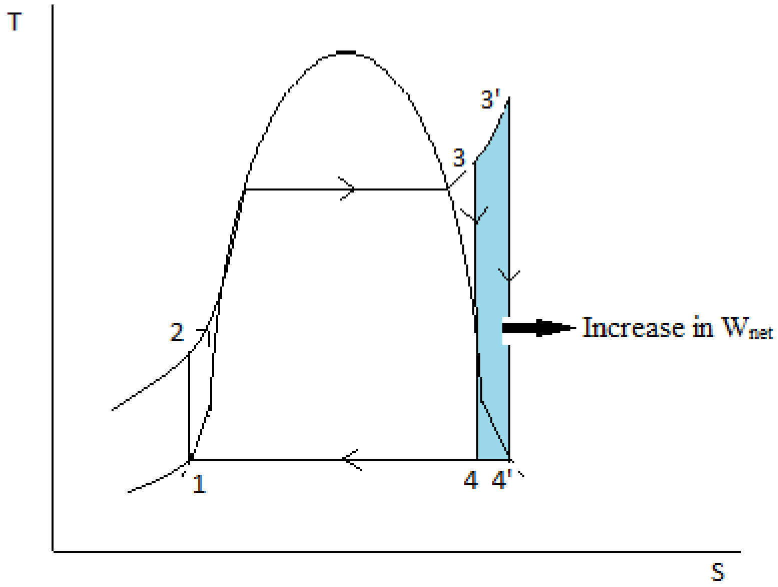


Figure C1: Effect of temperature on the efficiency of the Rankine cycle

The effect of temperature increase on the performance of the Rankine cycle is shown in Figure C1 on a T-s diagram. Due to the direct relationship between pressure and temperature, an increase in boiler operating pressure also raises the temperature of the steam. In Figure C1, the path 3-4 represents the part of the Rankine cycle in which there is isentropic expansion of steam in the turbine resulting in the production of work (W_{net}). As seen from Figure C1, a rise in steam temperature from 3 to 3' leads to an increase in W_{net} as represented by the shaded area. Since efficiency is a ratio of W_{net} to total heat input, the extra work produced results in an overall increase in efficiency of the cycle. Hence the increase in efficiency as observed with increasing temperature/operating pressure.

APPENDIX D1 – SUMMARRY OF ECONOMIC RESULTS

Table D.1.1: Summary of Economic Results of BMECP Models under Efficient Mill Conditions

		Combustion				Partial Fast Pyrolysis	Pure Fast pyrolysis
	Units	30bar BPST	40bar CEST	63bar CEST	82bar CEST		
INVESTMENT COST	M$						
Purchased Equipment		34.37	44.72	59.60	72.45	50.59	63.41
Other [a]		9.74	12.96	15.45	17.50	16.78	20.31
Indirect Cost [b]		6.99	9.16	11.86	14.18	10.71	13.27
Total TCI		51.1	66.84	86.91	104.13	78.08	97.00
Adjusted TCI [c]		**68.54**	**89.65**	**116.58**	**139.67**	**104.72**	**130.10**
OPERATING COST	M$						
Total Raw Material		18.14	18.14	18.14	18.14	18.14	18.14
Operating Labour and Maintenance		0.77	1.57	1.85	3.16	2.15	2.01
Operating Charges		0.15	0.31	0.35	0.66	0.43	0.39
Plant Overhead		0.39	0.78	0.93	1.58	1.08	1.01
G and A Expenses		1.56	1.66	1.70	1.88	1.74	1.72
Total Operating Cost		**21.01**	**22.47**	**22.98**	**25.43**	**23.54**	**23.27**
REVENUES	M$						
Main Product Sales		50.40	59.13	68.48	70.62	28.95	22.71
By-product sales		0.00	0.00	0.00	0.00	14.97	13.82
Total Product Sales		**50.40**	**59.13**	**68.48**	**70.62**	**43.92**	**36.53**
PRODUCTS YIELD PERFORMANCE							
Total power	MW	24.42	27.61	31.03	31.81	16.58	11.05
Net exportable power	kW/tch	61.40	72.04	83.44	86.02	35.28	16.85
Total HP steam	ton/h	161.73	159.56	153.15	150.59	128.72	123.73
Bio-oil	ton/h					9.36	16.69
Biochar	ton/h					2.32	0.00

[a] consist of cost for additional expenses and other support infrastructure such as painting, civil, electrical and instrumentation, insulation, equipment setting and freight and taxes charges.

[b] Total cost for G and A overheads, contractor fee and contingencies.

[c] Cost adjusted to the start date of basic engineering.

Table D.1.2: Summary of Economic Results of BMECP Models under Less Efficient Mill Conditions

		Combustion				Partial Fast Pyrolysis	Pure Fast Pyrolysis
	Units	30bar BPST	40bar CEST	63bar CEST	82bar CEST		
INVESTMENT COST	**M$**						
Purchased Equipment		33.79	43.52	59.59	72.12	51.20	67.86
Other [a]		9.64	12.72	15.37	17.42	16.96	21.30
Indirect Cost [b]		6.89	8.93	11.85	14.12	10.83	14.12
Total TCI		50.32	65.17	86.81	103.66	78.99	103.28
Adjusted TCI [c]		**67.50**	**89.39**	**116.43**	**139.03**	**105.94**	**138.53**
OPERATING COST	M$						
Total Raw Material		18.14	18.14	18.14	18.14	18.14	18.14
Operating Labour and Maintenance		0.82	1.82	1.81	1.85	2.14	2.03
Operating Charges		0.15	0.35	0.35	0.35	0.43	0.39
Plant Overhead		0.41	0.91	0.90	0.93	1.07	1.02
G and A Expenses		1.56	1.70	1.70	1.70	1.74	1.73
Total Operating Cost		**21.09**	**22.92**	**22.90**	**22.97**	**23.53**	**23.31**
REVENUES	M$						
Main Product Sales		42.13	52.70	62.1	67.03	40.44	25.42
By-product sales		0.00	0.00	0.00	0.00	9.97	16.55
Total Product Sales		**42.13**	**52.70**	**62.1**	**67.03**	**50.41**	**41.97**
PRODUCTS YEILD PERFORMANCE							
Total power	MW	21.40	25.26	28.68	30.50	20.78	15.29
Net exportable power	kW/tch	51.32	64.20	75.60	81.68	49.27	30.96
Total HP steam	ton/h	161.73	159.56	151.15	150.59	152.92	152.90
Bio-oil	ton/h					6.24	12.17
Biochar	ton/h					1.55	0.00

[a] consist of cost for additional expenses and other support infrastructure such as painting, civil, electrical and instrumentation, insulation, equipment setting and freight and taxes charges.

[b] Total cost for G and A overheads, contractor fee and contingencies.

[c] Cost adjusted to the start date of basic engineering.

APPENDIX D2 – BREAKDOWN OF TOTAL CAPITAL INVESTMEMT OF BMECP MODELS

Table D.2.1: Breakdown of Total Capital Investment for Combustion Based BMECP Models

		EFFICIENT MILL				LESS EFFICIENT MILL			
	Units	30 bar BPST	40 bar CEST	63 bar CEST	82 bar CEST	30 bar BPST	40 bar CEST	63 bar CEST	82 bar CEST
Combustion and steam generation	$	27650700.00	34219200.00	48820100.00	60862800.00	27591300.00	34219000.00	48810200.00	60862800.00
Turbine cycle and power production	$	5929300.00	9529300.00	10295100.00	10465200.00	5482400.00	8986100.00	9770200.00	10177700.00
	$	33580000.00	43748500.00	59115200.00	71328000.00	33073700.00	43205100.00	58580400.00	71040500.00
TCI	$	68543400.00	89645100.00	116579000	139674000.00	67500200.00	89388300.00	116428000.00	139032000.00
Specific TCI	$/MW	2806855.00	3246834.00	3756977.00	4390883.00	3154215.00	3538729.00	4059554.00	4558426.00

Table D.2.2: Breakdown of Total Capital Investment for Pyrolysis Based BMECP Models

		EFFICIENT MILL		LESS EFFICIENT MILL	
	Units	Pure Fast Pyrolysis	Partial Fast Pyrolysis	Pure Fast Pyrolysis	Partial Fast Pyrolysis
Feed preparation	$	10231200.00	4404700.00	10231200.00	3316300.00
Pyrolysis	$	7200400.00	3099900.00	7200400.00	2333900.00
Quench	$	6156400.00	3259000.00	6156400.00	2584800.00
Heat recovery and recycle	$	32289500.00	32079500.00	35508700.00	34356700.00
Oil recovery	$	512000.00	471800.00	512000.00	464700.00
Steam and power cycle	$	7076900.00	7264100.00	8216500.00	8217600.00
	$	63466400.00	50579000.00	67825200.00	51274000.00
TCI	$	130095000.00	104720000.00	138532000.00	105935000.00
Specific TCI	$/MW	11773303.00	6316043.00	9060301.00	5097930.70

APPENDIX D3 - SENSITIVITY ANALYIS TABLES

Table D.3.1: Response of BMECP Models to Changes in Bagasse and Electricity Prices (Efficient Mill)

		30 bar BPST		40bar CEST		63bar CEST		82bar CEST		Pure Fast Pyrolysis		Partial Fast Pyrolysis	
		NPV	IRR	NPV	IRR	NPV	IRR	NPV	IRR	NPV	IRR	NPV	IRR
Raw Material Cost													
% Change	$/dry ton												
-30	39.20	329.33	43.00	391.32	40.50	466.34	38.40	442.73	33.30	138.86	17.40	195.62	24.70
-10	50.40	285.25	39.00	347.25	37.40	422.27	36.00	398.62	31.30	85.00	14.80	151.66	21.70
0	56 .00	256.17	36.31	316.82	35.38	389.97	34.50	364.64	29.98	25.39	13.80	127.44	20.75
10	61.60	241.17	35.22	303.17	34.30	378.19	33.60	354.58	29.20	30.97	11.80	106.95	18.50
30	72.80	197.09	30.90	157.00	31.20	334.11	31.20	310.50	27.10	-23.26	8.50	61.97	15.20
Electricity Price													
% Change	$/kWh												
-30	0.174	95.94	21.40	129.24	21.80	173.24	22.20	142.36	18.70	2.21	10.10	32.87	12.90
-10	0.223	207.51	32.10	259.89	31.40	324.57	30.80	298.60	26.60	39.39	12.30	97.24	19.70
0	0.248	256.17	36.31	316.82	35.38	389.97	34.50	364.64	29.98	25.39	13.80	127.44	20.75
10	0.273	318.91	41.80	390.53	40.20	475.89	38.80	454.64	33.80	76.58	14.30	161.41	22.30
30	0.322	430.31	51.20	521.18	48.70	627.22	46.40	610.68	40.50	113.76	16.20	224.86	26.40

Table D.3.2: Response of BMECP Models to Changes in Bagasse and Electricity Prices (Less Efficient Mill)

		30 bar BPST		40bar CEST		63bar CEST		82bar CEST		Pure Fast Pyrolysis		Partial Fast Pyrolysis	
		NPV	IRR	NPV	IRR	NPV	IRR	NPV	IRR	NPV	IRR	NPV	IRR
Raw Material Cost													
% Change	$/dry ton												
-30	39.20	237.89	35.40	336.02	45.10	396.40	34.80	431.53	33.00	148.35	19.10	265.89	28.80
-10	50.40	193.81	31.20	291.96	40.90	352.33	32.40	387.46	30.90	104.24	16.60	221.87	26.10
0	56 .00	164.72	28.38	261.35	37.21	329.96	30.79	353.49	29.62	78.62	15.30	197.75	25.42
10	61.60	149.76	26.90	247.89	36.70	308.25	29.90	343.38	28.80	59.48	13.90	177.84	23.20
30	72.80	105.26	22.30	203.82	32.50	264.18	27.40	299.30	26.70	14.72	11.00	132.95	20.20
Electricity Price													
% Change	$/kWh												
-30	0.174	30.74	14.10	94.95	21.80	124.43	19.10	143.22	18.90	-2.72	9.80	65.56	15.4
-10	0.223	125.20	24.50	211.74	33.40	261.73	27.30	291.36	26.40	53.67	13.60	155.51	21.80
0	0.248	164.72	28.38	261.35	37.24	329.96	30.79	353.49	29.62	78.62	15.30	197.75	25.42
10	0.273	218.34	33.30	328.10	44.00	398.85	34.80	439.45	33.30	110.16	16.90	244.19	27.40
30	0.322	311.48	41.60	444.47	54.20	535.97	41.90	587.60	39.70	166.09	20.00	332.87	32.60