COMPARISONS OF THE TECHNICAL, FINANCIAL RISK AND LIFE CYCLE ASSESSMENTS OF VARIOUS PROCESSING OPTIONS OF SUGERCANE BAGASSE TO BIOFUELS IN SOUTH AFRICA

By

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ABSTRACT

Through many years of research, a number of production schemes have been developed for converting lignocellulosic biomass into transport fuels. These technologies have been assessed through a number of techno-economic studies for application in a particular context in terms of the technical and economic feasibility. However, previous studies using these methods have tended to lack vigour in various aspects. Either the energy efficiency of the processes were not maximised through adequate heat integration, or a competing technology which existed was not considered. From an economic perspective, the financial models would often lack the vigour to account for the risk and uncertainty that is inherent in the market prices of the commodities. This phenomenon is especially relevant for the biofuel industry that faces the full fledge of uncertainties experienced by the agricultural sector and the energy sector. Furthermore, from an environmental perspective, the techno-economic studies had often ignored the environmental impacts that are associated with biofuel production. Thus, a comparative study could have favoured an option due to its economic feasibility, while it could have had serious environmental consequences.

The aim of this study was to address these issues in a South African context, where biofuels could be produced from sugarcane bagasse. The first step would be to modify an existing simulation model for a bioethanol scenario that operates with a Separate Hydrolysis and Fermentation (SHF process) configuration into a second processing scenario that operates with a Simultaneous Saccharification and Fermentation (SSF process) configuration using reliable experimental data. The second step was to ensure that the maximum energy efficiency of each scenario was realised by carrying out pinch point analysis as a heat integration step. In contrast to these biological models is the thermochemical model that converts bagasse to gasoline and diesel via gasification, Fischer-Tropsch synthesis and refining (GFT process). While there were no significant advances in technology concerning this type of process, the energy efficiency was to be maximised with pinch point analysis.

The GFT process obtained the highest energy efficiency of 50.6%. Without the affects of pinch point technology, the efficiency dropped to 46%, which thus emphasises the importance of heat integration. The SSF had an efficiency of 42.8%, which was superior to that of the SHF at 39.3%. This resulted from a higher conversion of biomass to ethanol in the SSF scenario. Comparing the SHF model to an identical model found in literature that did not have pinch point retrofits, this study showed lower efficiency. This arose because the previous study did not account for the energy demands of the cold utility systems such as the cooling tower operation, which has been shown in this study to account for 40% of the electrical energy needs.

The economic viability of all three processes was assessed with Monte Carlo Simulations to account for the risks that the fluctuations in commodity prices and financial indices pose. This was accomplished by projecting the fluctuations of these parameters from samples of a historical database that has been transformed into a probability distribution function. The consequences were measured in terms of the Net Present Value (NPV) and Internal Rate of Return (IRR) for a large number of simulations. The results of these variables were aggregated and were then assessed by testing the probability that the NPV<0, and that the IRR recedes below the interest rate of 12.64%. The investment was thus deemed unfeasible if these probabilities were greater than 20%.

Both biological models were deemed profitable in terms of this standard. The probabilities were 13% for the SSF and 14% for the SHF. The GFT process however was deemed completely unfeasible because the probability that the $NPV \leq 0$ was 78%. Given that the GFT process had the highest energy efficiency, this result arises mainly because the capital investment of 140,000USD/MW_{HHV} of biomass energy input is to enormous for any payback to be expected.

The environmental footprint of each process was measured using Life Cycle Assessments (LCAs). LCAs are a scientifically intricate way of quantifying and qualifying the effects of a product or process within a specified boundary. The impacts are assessed on a range of environmental issues, such as Global Warming, Acidification, Eutrophication and Human toxicity. Furthermore, if the project under concern has multiple output products, then the impacts are distributed between the output products in proportion to the revenue that each generates.

The impacts were either relative to the flow of feedstock, which was 600MW of bagasse, or to the functional unit, which was the amount of fuel required to power a standard vehicle for a distance of 1 kilometre. In either case, the GFT scenario was the least burdening on the environmental. This was expected because the GFT process had the highest energy efficiency and the process itself lacked the use of processing chemicals. Relative to the feedstock flow, the SSF was the most environmentally burdening scenario due to the intensive use of processing chemicals. Relative to the functional unit, the SHF was the most severe due to its low energy efficiency.

Thus, the following conclusions were drawn from the study:

- The GFT is the most energy and environmentally efficient process, but it showed no sign of economic feasibility.

- There is no significant difference in the economic and environmental evaluation of the SSF and SHF process, even though the SSF is considered to be a newer and more efficient process. The major cause of this is because the setup of the SSF model was not optimised.

OPSOMMING

Deur baie jare van navorsing is ‘n aantal produksie-skemas vir die omskakeling van lignosellulose biomassa na vloeibarebrandstof ontwikkel. Hierdie tegnologië is geassesseer ten opsigte van die tegniese en ekonomiese haalbaarheid deur middel van tegno-ekonomiese studies in bepaalde tekste. Tog het hierdie vorige studies besliste beperkings gehad. Of die energie-doeltreffendheid van die proses is nie gemaksimeer deur voldoende hitte-integrasie nie, of 'n mededingende tegnologie wat bestaan is nie oorweeg nie. Vanuit 'n ekonomiese perspektief, was die finansiële modelle dikwels nie die omvattend genoeg om rekening te hou met die risiko en onsekerheid wat inherent is in die markpryse van die kommoditeite nie. Hierdie verskynsel is veral relevant vir die biobrandstof bedryf wat die volle omvang van onsekerhede ervaar waaraan die landbousektor en die energiesektoronderhewig is. Verder het die tegno-ekonomiese studies dikwels die omgewingsimpakte wat verband hou met biobrandstofproduksie geïgnoreer. Dus kon ‘n opsie deur die ekonomiese haalbaarheid bevoordeel word, ten spyte van die ernstige omgewingsimpakte wat dit kon inhou.

Die doel van hierdie studie was om hierdie kwessies aan te spreek in 'n Suid-Afrikaanse konteks, waar biobrandstof uit suikerriet bagasse geproduseer kan word. Die eerste stap was om 'n bestaande simulasiemodel vir 'n bio-scenario wat met Afsonderlike Hidroliese en Fermentasie (SHF proses) stappe werk, te modifiseer vir 'n tweede verwerking scenario wat met 'n gelyktydige Versuikering en Fermentasie (SSF proses) konfigurasie werk. Die verandering is gedoen deur die gebruik van betroubare eksperimentele data. Die tweede stap was om te verseker dat elke scenario die maksimum energie-doeltreffendheid het, deur 'n hitte-integrasie stap, wat gebruik maak van “pinch-point” analise. In teenstelling met hierdie biologiese modelle, is daar die thermochemiese roete waar petrol en diesel van bagasse vervaardig word via vergassing, Fischer-Tropsch-sintese en rafinering (GFT proses). Daar was geen betekenisvolle vooruitgang in tegnologie vir hierdie proses nie, maar die energie-doeltreffendheid is gemaksimeer word deur energie-integrasie.

Die GFT proses toon die hoogste energie-doeltreffendheid van 50,6%. Sonder die invloed van energie-integrasie het die doeltreffendheid gedaal tot 46%, wat dus die belangrikheid van hitte-integrasie beklemtoon. Die SSF het 'n effektiwiteit van 42,8% gehad, wat beter was as dié 39,3% van die SHF opsie. Hierdie hoër effektiwiteit wasas gevolg van die hoër omskakeling van biomassa na etanol in die SSF scenario. Die energie doeltreffendheid vir die SHF-model was laer as met 'n identiese model (sonder energie-integrasie) wat in die literatuur gevind wat is. Dit het ontstaan

omdat die vorige studie nie 'n volledig voorsiening gemaak het met die energie-eise van die verkillingstelselsnie, wat tot 40% van die elektriese energie behoeftes kan uitmaak.

Die ekonomiese lewensvatbaarheid van al drie prosesse is bepaal met Monte Carlo simulasies om die risiko's wat die fluktuasies in kommoditeitspryse en finansiële indekse inhou, in berekening te bring. Hierdie is bereik deur die projeksie van die fluktuasies van hierdie parameters aan die hand van 'n historiese databasis wat omskep is in 'n waarskynlikheid verspreiding funksie. Die gevolge is gemeet in terme van die netto huidige waarde (NHW) en Interne Opbrengskoers (IOK) vir 'n groot aantal simulasies. Die resultate van hierdie veranderlikes is saamgevoeg en daarna, deur die toets van die waarskynlikheid dat die NPV <0, en dat die IRR laer as die rentekoers van 12,64% daal, beoordeel. Die belegging is dus nie realiseerbaar geag as die waarskynlikhede meer as 20% was nie. Beide biologieseprosesse kan as winsgewend beskou word in terme van bostaande norme. Die waarskynlikhede was 13% vir die SSF en 14% vir die SHF. Aangesien die NHW van die GFT-proses onder 0 met ‘n waarskynlikheid van 78% is, is die opsie as nie-winsgewend beskou. Gegewe dat die GFT-proses die hoogste energie-doeltreffendheid het, is die resultaat hoofsaaklik omdat die kapitale belegging van 140,000 USD / MWHHV-biomassa energie-inset te groot is, om enige terugbetaling te verwag.

Die omgewingsvoetspoor van elke proses is bepaal deur die gebruik van Lewens Siklus Analises (“Life Cycle Assessments”) (LCAS). LCAS is 'n wetenskaplike metodeom die effek van ‘n produk of proses binne bepaalde grense beide kwalitatief en kwantitatief te bepaal. Die impakte word beoordeel vir 'n verskeidenheid van omgewingskwessies, soos aardverwarming, versuring, eutrofikasie en menslike toksisiteit. Voorts, indien die projek onder die saak verskeie afvoer produkte het, word die impakte tussen die afvoer produkte verdeel, in verhouding tot die inkomste wat elkeen genereer.

Die impak was met of relatief tot die vloei van roumateriaal (600MW van bagasse), of tot die funksionele eenheid, wat die hoeveelheid van brandstof is om 'n standaard voertuig aan te dryf oor 'n afstand van 1 kilometer. In al die gevalle het die GFT scenario die laagste belading op die omgewing geplaas. Hierdie is te verwagte omdat die GFT proses die hoogste energie-doeltreffendheid het en die proses self nie enige addisionele chemikalieë vereis nie. Relatief tot die roumateriaal vloei, het die SSF die grootse belading op die omgewing geplaas as gevolg van die intensiewe gebruik van verwerkte chemikalieë. Relatief tot die funksionele eenheid, was die SHF die swakste as gevolg van sy lae energie-doeltreffendheid.

Dus, kan die volgende gevolgtrekkings gemaak word:

- Die GFT is die mees energie- en die omgewings doeltreffende proses, maar is nie ekonomies lewensvatbaar nie.
- Daar is geen beduidende verskil in die ekonomiese en omgewings evaluering tussen die SSF en SHF proses nie, selfs al word die SSF proses beskou word as die nuwer en meer doeltreffende proses. Die grootste oorsaak hiervan was omdat die SSF-model nie geoptimeer is nie.

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NOMENCLATURE

GFT - Gasification and Fisher-Tropsch Synthesis

SSCF - Simultaneous Saccharification and Co-Fermentation Reactor

SHCF - Separate Hydrolysis and Co-Fermentation Reactors

SSF process/scenario - A complete process design scenario incorporating SSCF

SHF process/scenario - A complete process design scenario incorporating SHCF

NREL - National Renewable Energy Laboratory

P. Stipitis - *Pichia stipitis*

S. Cerevisae - *Saccharomyces cerevisiae*

IRR - Internal Rate of Return

NPV - Net Present Value

BOP - Balance of Plant

TDC - Total Direct Costs

TCI - Total Investment Costs

PPI - Producers Purchases Index

IDC - Industrial Development Corporation

SVO - Spent Vegetable Oil

PEA - Process Environmental Analysis

UNFCCC - United Nations Framework Convention on Climate Change

IEA - International Energy Agency

EPA-USA - Environmental Protection Agency of the United States of America

CTL - Coal to liquid fuel

BTL - Biomass to liquid fuel

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1 INTRODUCTION

1.1 BACKGROUND TO INVESTIGATION

The concept of a bio-refinery has been receiving an increasing amount of attention in the South African context in recent years, due to the promising potential that bio-fuels have as an alternative energy source to fossil fuels and to serve as a renewable energy source. However, the use of agricultural land to grow feedstock for biofuel production has come under severe criticism, due to the impact it potentially poses on the food industry (Naylor *et al*, (2010)). In addition, the cost of dedicated energy crops can render the biofuel production process unattractive from an economic point of view.

Agricultural wastes offer a much better alternative as a biomass feedstock for biofuel production. In South Africa, the maize, wheat and sugar industries produce high amounts of wastes. The sugar industry produces ±8million tonnes of bagasse per annum (Leibbrandt 2010), a proportion of which is burnt in boilers to produce the energy needs of sugar mills, leaving excess quantities available for alternative products. Biofuels is an energy product that serves as a very attractive alternative for which the excess bagasse can serve as feedstock, due to its potential to reduce the global warming effects caused by fossil fuel combustion. The viability would, as per usual, depend on a variety of factors, such as the state of technology, market situation, infrastructure and environmental friendliness of the product throughout its life cycle. Furthermore, socio-economic development potentials are also major incentives for investments, especially in a developing context.

Investment decisions depend mostly on the technical and economic feasibility, because investors would need to to know with an absolute degree of certainty that there will be a return on the investment. Environmental regulations on the other hand, such as those imposed by the National Environmental Management Act (NEMA), requires a pre-auditing procedure of the proposed facility to assess the potential environmental burden. Operating licenses will only be granted once certain criteria regarding environmental preservation are met.

A techno-economic model is a typical package that contains detailed process simulations and financial models to evaluate and compare the technical and economic feasibility of processing routes, which have been proven to obtain the processing objective. Such techniques have been previously used on processes that convert lignocellulosic biomass to biofuels, such as Anex *et al*, (2010). In a local context, Leibbrandt (2010) used this type of evaluation for processes that convert bagasse to biofuels. Both studies contained processing routes which were either thermochemical or biological. The technologies were proven at some point to convert cellulosic biomass to biofuels. Many laboratory-scale experiments have proven these capabilities with various technologies such as fermentation, pyrolysis and gasification-Fischer-Tropsch. Aden et al. (2002) has conducted pilot plant runs to prove that biological conversion of cellulosic biomass to ethanol is technically feasible.

In these studies, a design for optimal energy usage for the processes under consideration was not carried out with the necessary detail that is required to reach a configuration for optimised energy usage. This does not necessarily mean that none of the excess heats were used, but the method of doing so was not in accordance with any of the standard heat integration techniques, such as Pinch Point Analysis. The application of such techniques has been proven to be a critical component in achieving the maximum energy efficiency of a process, and ultimately, a profitable investment, because excessive expenditure on obtaining energy utilities was abated.

While previous studies were conclusive on the economic and technical status of the processing scenarios were compared, the economic models were static in nature, i.e. deterministic. Thus, economic parameters that are known to have adverse fluctuations, due to supply and demand, etc, were treated as static averages. Agricultural industries have established economic evaluative methods to account for risks involved with such fluctuations (Richardson et al. 2000). These methods are called Financial Risk Assessments (also known as Probabilistic Economic Models) and they quantify the probability of a successful outcome of a proposed financial undertaking. This is achieved by intrinsically including the fluctuations in commodity prices and economic indices through characterising their inputs with probability distributions. The biofuel industry, being a derivative of the agricultural industry, is in essence an “agri-business” and should be economically evaluated using the same financial risk assessment techniques that are applied to the agricultural industry. Such evaluations have been carried for biofuel industries in a South African context (Richardson et al. 2007). The biofuel industry is also a derivative of the energy sector, and would also be severely affected by fluctuations in the prices of energy commodities.

The question pertaining to environmental sustainability was not regarded for the studies of Leibbrandt (2010), Aden et al. (2002) or Richardson et al. (2007). An environmental auditing study by Sheehan et al. (2004) however, complimented the study of Aden et al. (2002) at a later stage. While Environmental Impact Assessments (EIA) are commonly used to audit the potential environmental impact of a proposed operation, the results are often based on qualitative interpretations of haphazardly collected data that is seemingly unrelated in a checklist form. Life Cycle Assessment studies though, offer a far more technical and comprehensive form of environmental auditing, and would be more appropriate and accurate in describing the environmental pros and cons of a proposed biofuel facility, regardless of the context.

A geographical setting of a biofuel facility would pose implications on the technical, economic and environmental assessments. That is because the type and usage of natural resources, infrastructure and the response of an ecosystem to environmental impacts varies with the geographical features of varying localities. The region of the sugar producing industry in Southern Africa is on the coastline of Northern KwaZulu-Natal and Southern Mozambique. Therefore, a proposed biofuel plant that sources its feedstock as excess bagasse should be centrally located in this region to the sugar mills that produce the waste bagasse, as proposed by Leibbrandt (2010). There is enough bagasse produced in this region to supply a central biofuel facility that has a biomass thermal input of 600MW. Such a facility has the capacity to process 909474 tons of dry bagasse per annum.

1.2 AIMS AND OBJECTIVES

Thermochemical and biological processing models for evaluating and comparing process routes for converting bagasse to biofuels in a South African context have previously been completed in techno-economic studies, such as that of Leibbrandt (2010). However, the concerns with this, as is with some others, such as Anex *et al*, (2010), are raised:

1. The state of technology that the biological processes were based on is not currently up to date, as it was based on the NREL 2002 designs (Aden et al. 2002) for ethanol production through fermentation. On the other hand, thermochemical process models for gasification of biomass followed by FT synthesis of gasoline and diesel developed were based on developing technology with additional optimisation.
2. A thorough investigation for heat integration to mitigate expenses was not carried out for either the biological or thermochemical processes. Thus, the reported energy efficiencies are not reflective of the optimal.
3. The comparisons were carried out using economic models that were static and did not intrinsically account for the range of risks and uncertainties that are typically experienced in the agricultural and energy sectors.

4. Being purely techno-economic studies, the environmental consequences were not considered as they were excluded from the scope of the project.

Thus, in consideration of these four concerns, the aims and objectives of the ensuing study are as follows:

1. To update the biological process models developed to the most recent state of technology and also take into account the implications of locality on technology for all process models.
2. To carry out extensive heat integration exercises on each process using pinch point analysis.
3. To consider all processes, from the base biological case ((Leibbrandt 2010), Aden et al. 2002)) along with the technological updates considered, and the thermochemical route, for financial risk assessment using Monte Carlo Simulations.
4. Assess and compare the environmental burdens of the biological route to that of the thermochemical route while considering the implications of the local geography.

1.3 HYPOTHESIS

Regarding the aims mentioned above, the following hypotheses are made:

1. The upgrading of a biological scenario to a more recent technological standing to promote biofuel production will greatly enhance the technical feasibility, as the liquid fuel output will increase and capital requirements will reduce.
2. Pinch point retrofits will enhance the energy efficiencies as the usage of waste heats to satisfy heat burdens will diminish the energy utility requirements.
3. The GFT thermochemical scenario would maintain its superior energy efficiency status over the biological models because the biological models inherit energy intenseness by the vast amounts of water that is involved in the processing.
4. The probabilistic economic assessment of the biological process will show more viability, as it is not expected that the enormous capital requirement for the GFT process would be met even if the energy efficiency was improved by pinch point analysis. The Internal Rate of Return (IRR) for this option according to Leibbrandt (2010) is only around 10%, and this would probably need to double in order to produce a payback on the investment that is greater than that of the optimised biological model.
5. There will be a trade-off between environmental burdening of the two differing processing options. While the impacts associated with the product of the biological production process

have a lesser burdening potential than the GFT product, the process itself can impose severe environmental burdening.

1.4 PROJECT WORKFLOW

The order of the tasks that are described in the Aims and Objectives section were designed to arrive at a set of conclusions that were relative to the set of Hypothesis. The order of tasks is represented in Figure 1. Each phase of this project though has its own technical and scientific detail that is seemingly independent of the others (besides phase 1 and 2 that are related). As an example, the section on technological retrofits concentrates on the sciences of bio-processing (amongst others), while that of Life Cycle Assessments concentrates on the formulation of a scientific method to evaluate impacts associated with the life of a product or its process.

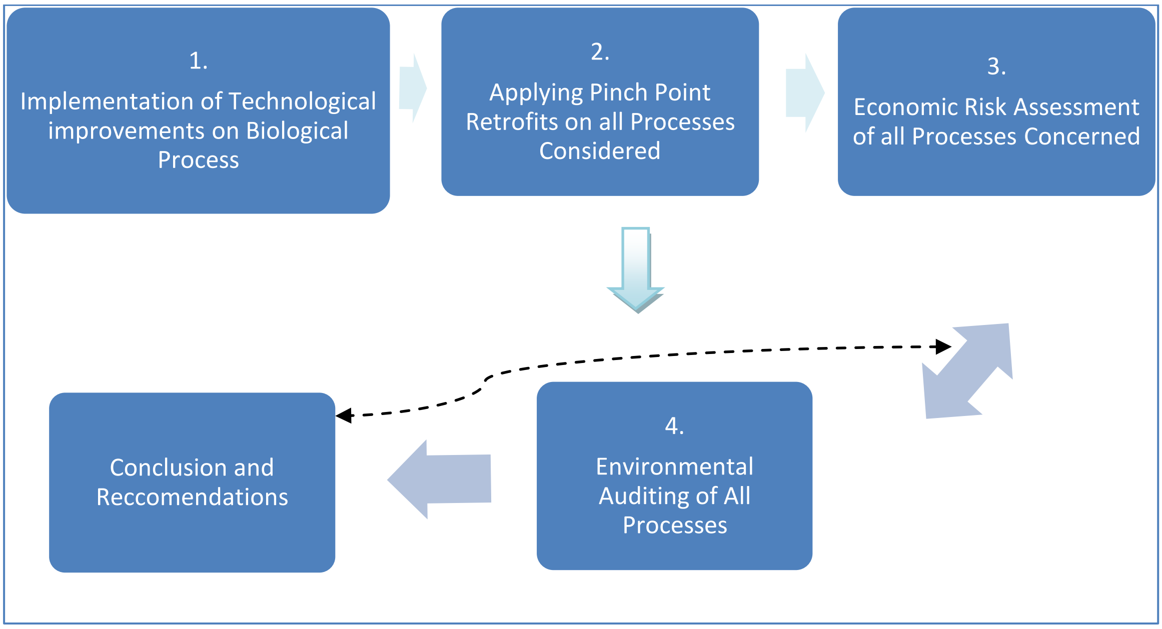

Figure 1: Workflow

Thus, the thesis will contain three sections: Section A containing phase 1 and 2, Section B containing phase 3, and Section C containing phase 4. Each section will contain a review of literature that is relevant to the methodology and analysis of the technical data. The literature of each project section is used to formulate the scientific methodology that is to be used in that section, and does not affect the methodology in other sections. This does NOT imply in the least that the sections of the project are independent. Section A is a technological development phase, and decisions there in will adversely affect the outcomes of Section B and Section C. Section B and C on the other hand are evaluative sections and they weigh against or complement each other. However, if the literatures of all the sections are conglomerated in a single chapter, then it will not be coherent.

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2 PROCESS ADVANCEMENTS AND PINCH POINT TECHNOLOGY

2.1 LITERATURE REVIEW: PROCESS MODIFICATIONS AND IMPROVMENTS

2.1.1 Overview of Current Process

A process model for simulating the conversion of lignocellulosic biomass from cornstover to fuel grade ethanol was reported by the NREL design report of Aden *et al.* (2002). This model formed the skeletal framework of subsequent techno-economic studies regarding second-generation ethanol production, such as Kazi *et al.* (2010) who also considered cornstover as feedstock, and of Leibbrandt (2010) who adapted the process for sugarcane bagasse. A generic block flow diagram for a biological process for converting biomass to ethanol is presented in Figure 2 below.

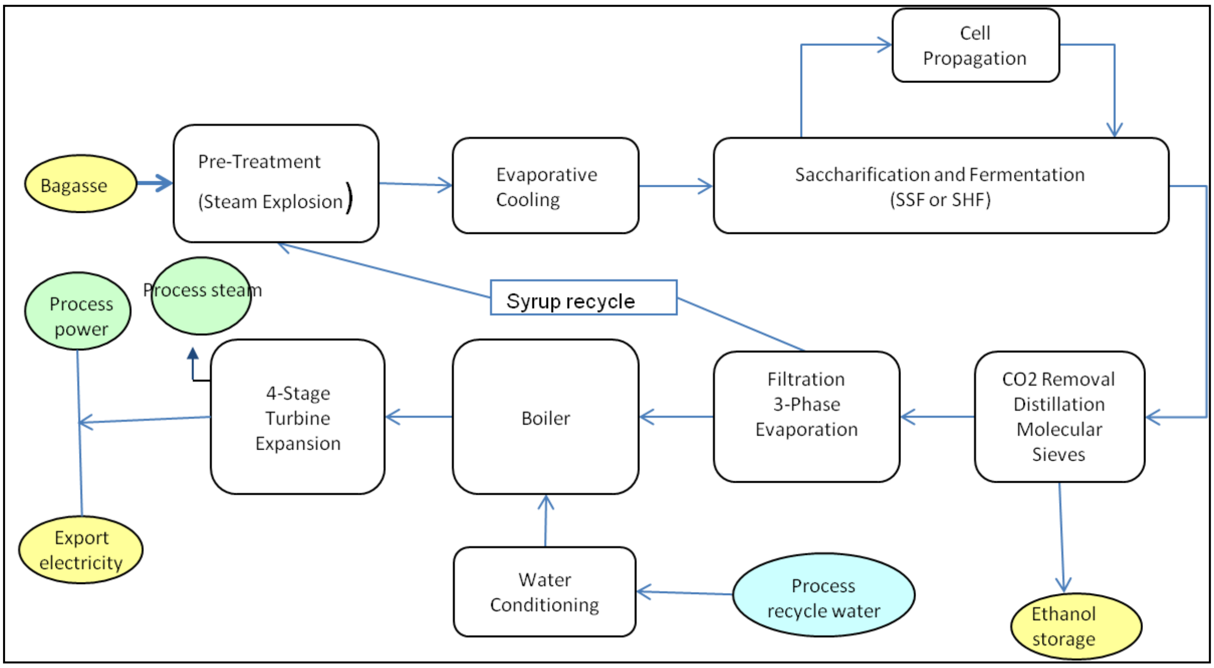


Figure 2: Generic Biological Process

The bagasse received from a sugar-mill plant is fed to a pre-treatment reactor, where the digestibility and porosity of the biomass is increased. Several methods for pre-treatment, such as steam explosion catalysed with dilute acid, liquid ammonia under high pressure and liquid hot water can be considered. Of these options, the steam explosion technique is the most viable from energy efficiency and an economic point of view because the process can occur at solid concentrations of up to 50% solids. This has been observed by Kazi *et al.* (2010) and Leibbrandt (2010). The technical drawback of this option is the severity of process generates degradation products such as furfural (Sendelius 2005), that inhibits the microbial activity downstream, are yielded from such treatment methods. This problem could be negated by using a yeast strain that was engineered to withstand the high levels of inhibitors (Martın *et al.* (2006)). Yeast cells are typically accustomised to toxic environments by the gradual introduction the inhibitory chemicals into the cultivation schemes of yeast cells (Martın *et al.* (2002), Ohgren *et al.* (2006), Aden *et al.* (2002) and many others).

The treated pulp is then fed to a hydrolysis reactor where it undergoes saccharification. In this process, the cellulose and hemicelluloses is converted to glucose through enzymatic hydrolysis. A bleed stream from the hydrolysate containing inhibitory chemicals is diverted to a seed reactor train where yeast is cultivated for the primary fermentation vessel (Aden *et al.* 2002). Nutrients and 35% excess oxygen is supplied to the reaction vessel to facilitate further growth of yeast.

The design specification for the hydrolysis of glucan of a pulp resulting from sulphuric acid (H_2SO_4) catalysed steam explosion, in the study of Leibbrandt (2010) was 83%. Thus, any technological improvement on steam explosion would have to meet or improve on this specification in order to be considered. The hydrolysate is pumped to the fermentation train were sugars are converted to ethanol with a modified yeast. The conversion from glucose to ethanol was 88% while that of xylose was 44% based on the experimental work of Martın *et al.* (2006). Thus, the overall conversion from glucan to ethanol was 79.2% for a setup separating saccharification and fermentation (i.e. SHF), considering that there was hydrolysis of cellulose 7%, in addition to the enzymatic hydrolysis of 83% and the ethanol conversion of 83%.

The fermenting tank would require cooling to maintain a suitable temperature that is suitable for the organisms. This would also be true for the seeding tanks. An optimum temperature for fermentation has been reported in literature (Martın *et al.* (2002), Ohgren *et al.* (2006)) to be 30^{o}C. Considering that the cooling water for the facility would be sourced from the nearby Indian Ocean, where sea water temperatures are likely to reach 28^{o}C in summer (www.sea-temperature.com), refrigeration systems would be the likely mode of cooling (plant situated in the coastal strip of Kwazulu-Natal (Leibbrandt (2010))). This is because a temperature deference of 2^{o}C between the fermentation reactor contents and sea water would not serve as an adequate driving force for heat transfer, given that absolute minimum temperature recommended by March (1998) was 3^{o}C. Even this extreme minimum was only recommended at very cold temperatures where the Coefficient of Performance (COP) of the refrigeration system drops below 1.

The fermentation product, which is known as beer is refined in the ethanol recovery section where carbon dioxide (CO_2), water and lignin are separated from the beer to produce an ethanol product of 99.6% at a recovery of 99.7%.. The specifications of this section of the plant were taken from McAloon *et al.* (2000) for a second generation ethanol plant utilizing cornstover as a feedstock. These specifications are realistic estimations of ethanol purification from fermenting reacting effluents which is the beer. The CO_2 is removed in two flash drums, and any loss of ethanol is recovered from the vapour in the scrubber. The liquid effluent from the scrubber is fed with the liquid still of the flash drum to the beer column. The beer column separates lignin and any other heavy residues to produce a 55% ethanol distillate, which is purified to 95% in a rectifying column and ultimately 99.6% in the molecular sieves. A 0.3% loss of ethanol to the residues occurs in the entire recovery section.

The residue undergoes an initial evaporation step, where 25-30% of the water is evaporated, followed by a filtration in a pneumatic press, where the insoluble solid residues are removed from the syrup. From this syrup, 25% is split to the pre-treatment section to improve the overall ethanol conversion (Aden *et al.* (2002), Leibbrandt (2010)). The remaining liquid undergoes further concentration in a multiple effect evaporator which is then combined with the solid residue to power the boiler. Energy from the boiler provides the energy needs of the plant. The boiler produces saturated steam at 86bar (at $\pm$ 750^{o}C) which provides thermal energy where needed and passes through steam turbines to generate electricity. Rudimentary heat integration was employed throughout the model to maximize energy efficiency.

2.1.2 Technological Advances

Newer consideration for steam explosion has been focused on sulphur dioxide (SO_2) impregnation of bagasse as an alternative to dilute H_2SO_4 for the catalysis of the pre-treatment stage. Experimental work carried out by Sendelius (2005) which compared dilute acid catalysis to SO_2 catalysis showed that xylose yields from xylan were higher in the case of SO_2 impregnation.

While the specification of glucan-glucose conversion from Leibbrandt (2010) was 83%, (Carrasco *et al.* 2009) obtained enzymatic hydrolysis of glucan to glucose on pulps treated with SO_2 impregnation of up to 87%. Thus, this data shows that technical feasibility of SO_2 impregnation in terms of the enzymatic performance downstream is at least equal with H_2SO_4 catalysis. This outcome was also confirmed with the outcomes of experiments performed by Sendelius (2005) and by Wyman *et al.*

(2011) (CAFI). Wyman *et al.* (2011) however, showed that the pre-treatment of biomass with lime caused a significantly better performance of the enzymatic hydrolysis downstream. This outcome is compromised though, since lime pre-treated biomass had low xylose yields. Furthermore, lime pre-treatment leads to additional problems associated with neutralization of the alkaline slurry with mineral acids.

It stands to reason, that the amounts of inhibitors formed through SO_2 impregnation would be lower than that of H_2SO_4 catalysis, since the pH of sulphorous acid is lower and the resulting pre-treatment conditions is less severe. This theoretical outcome is substantiated with the empirical evidence of the experimental work in Carrasco *et al.* (2009).

Steam explosion can have also have the desired effects of pre-treatment on lignocellulosic biomass without the prior SO_2 impregnation (Martın *et al.* 2002). Such an option would have desirable effects on the environmental impact as the absence of SO_2 in the system will diminish the potential to produce acid rain from the system's emissions. The xylose yields from such a setup, however, have been shown by Martın *et al.* (2002) to be 25% lower than the setup that uses SO_2 impregnated material. This would in turn reduce the ethanol yield by about 9%, considering that the content of xylan is +/- 36% of the carbohydrates in bagasse.

Simultaneous Saccharification and Fermentation (SSF) is often recognized as being advantageous over SHF because of the following reasons:

- reduced end product inhibition effects – in a hydrolysis reactor, the hydrolysis reaction is inhibited by the increasing levels of glucose (Ghosh *et al.* 1982). In a SSF reactor, these effects are minimized due to the simultaneous consumption of glucose.
- reduced cost of capital – since an SSF configuration combines the fermentation an hydrolysis reactions, there would only be capital expenditure one set of reactors instead of two (Aden and Foust 2009).

SSF overall conversions have been shown in literature to exceed the specifications of the SHF process in Leibbrandt (2010), in terms of glucan-ethanol and xylan-ethanol conversions. Sendelius (2005) experimented with steam exploded bagasse across a SSF unit to determine the yields of ethanol. The SSF test results showed that the theoretical conversions of cellulose achieved across the SSF step is about 80%, provided that the steam explosion conditions are correct, which is generally, that the temperature of the steam explosion should not exceed 205°C, else excessive amounts of inhibitors will form.

The conversion (Sendelius 2005) was only based on the cellulose though, and not on the combined amount of cellulose and hemicelluloses entering the pre-treatment step. The yeast in the SSF tests, which was the *Saccharomyces cerevisiae* (S. cerevisisae) native strain, has only significant fermentation rates for hexose sugars. These strains have very negligible uptakes of xylose, and thus no significant co-fermentation would take place.

Previous experimental work has shown that genetically advanced strains are capable of achieving high yields of ethanol based on the co-fermentation of hexose and pentose sugars. Rudolf *et al.* (2007) carried out SSCF experimental work with two co-fermenting strains, with the aim of establishing which organism had the highest ethanol yield from the xylose and cellulose entering the SSCF reactor. The strains that were compared were *P. stipitis* CBS6054, a natural xylose fermenting yeast, and *S. cerevisiae* TMB400, a genetically engineered strain capable of fermenting xylose. This engineered strain, however, also converts small amounts of the xylose to xylitol, which results in a decrease in the overall ethanol yield.

Sugarcane bagasse was pretreated with steam explosion catalysed with SO_2 impregnation at conditions similar to the optimums shown by Sendelius (2005), in that the maximum temperature of 205°C was not exceeded. The concentration of inhibitors in the hydrolysate of steam explosion by Rudolf *et al.* (2007) were much lower compared to the H_2SO_4 catalysed hydrolysate used in the model of Leibbrandt (2010).

The S. cerevisiae TMB400 produced better yields than the P. stipitis under all the experimental conditions considered. At a solids concentration of 5%, the conversion of ethanol by TMB400 from hemicelluloses and cellulose was 68%. The saccharification and fermentation in Leibbrandt (2010) was based on batch experiments where the water insoluble solids (WIS) content was 5% (Martın *et al.* (2006)). Even so, the design slurry density in the simulation models was 13% for hydrolysis because pilot plant data from Aden *et al.* (2002) confirmed that continuous operations can tolerate much higher slurry densities than the corresponding batch operation. The same rationale could subsequently be applied to data obtained from SSF batch processes so that similar ethanol concentrations in the beer products of SSF-SHF counterparts are achievable.

During the latter part of this study, the NREL published an updated design report (Humbird *et al.* (2011)), and thus the advances in that design over the NREL 2002 design were not considered in the current study. Either way, the major advances in that study were not directly applicable in the scope of this study. Firstly, the advancements pertained to the finer design details of a dilute acid pre-treatment setup, where as this study had by now assumed that SO_2 catalysed pre-treatment setup be used, due to its superior characteristics that was validated by experimental literature. Secondly, a variation in the configuration of a SHF system with batch reactors was considered. Such detail was not inclusive in the scope of this study, which focussed on the comparison of SHF and SSF processes on a continuous basis. Thirdly, the water treatment section differed drastically, as a consequence of the conditioning of the pre-treatment slurry with ammonia. That water treatment plant used biological treatment with digesters, which had shown to incur a tremendous in capital expenditure over the equivalent in the NREL 2002 report, which was the 3-phase evaporation scheme.

2.1.3 Overview of the Gasification-Fischer Tropsch process (GFT Process)

Kreutz *et al.* (2008). et al published a process model for simulating the conversion of biomass to Fischer-Tropsch transportation fuels via gasification, Fischer-Tropsch synthesis and petrochemical refining. Similar models were published by Liu *et al.* (2011) and Hamelinck *et al.* (2003). The process model of Kreutz *et al.* (2008) which used mixed prairie grass as a feedstock was subsequently adapted by Leibbrandt (2010) to utilize sugarcane bagasse. A schematic diagram of the process is given in Figure 3:

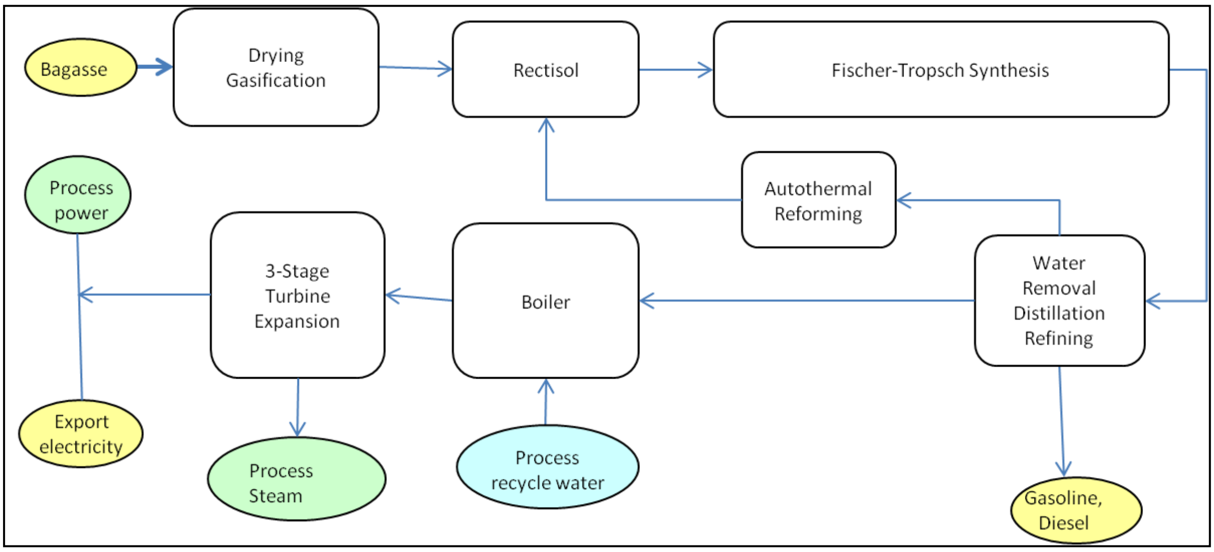


Figure 3: Gasification Fischer Tropsch Process (GFT)

The feedstock was initially dried, since the moisture content needed to be below 10% for the feedstock to be suitable for gasification. The gasification is critical to the feasibility of a process for converting biomass to transportation fuels because of the intensive energy usage, and because the quality of the resulting syngas composition, in terms of the molar ratio of carbon monoxide to CO_2, must be maximized to ensure an adequate conversion to the liquid fuels. Thus, an extensive optimization study was conducted in Leibbrandt (2010) over the gasification section using equilibrium modeling as a treatment response to varying inputs for operating conditions. Thus, the optimized operating conditions for the gasification, as yielded by the equilibrium modeling in that study, are given in Table 1.

Table 1: Gasification Conditions

Operating Condition	Value
Moisture [wt%]	5.00
Steam Biomass Ratio [wt/wt]	0.75
Temperature	1100K
Pressure	1 bar
O_2 Added/Stoichiometric Combustion O_2	0.25

After gasification, the particulates were removed with a cyclone. There were no units for tar cracking, since the gasifier was designed with a nickel catalyst to promote internal cracking (Kreutz *et al.* (2008)). The syngas was cooled to 350°C, and the heat from this was used generate steam at 510°C, 1atm for the energy needs of the gasification section. Any particulates that might have formed by condensation are captured in a ceramic filter. The syngas is further cooled to 40°C to facilitate the acid gas removal in the Rectisol unit, which operates at 2bar and 27°C, using methanol as a solvent for the acid gasses. The vast amounts of heat energy released by the cooling processes are used for feedstock drying (Hamelinck *et al.* 2003). All acid gasses that were removed are vented. The gasses could alternatively be sequestrated through geological injection to reduce the emissions of greenhouse gasses. Kreutz *et al.* (2008) has shown though, that such an action was technically and economically unfeasible because it caused a capital increase of 67 million USD (2007 based currency) and a decrease in the overall energy efficiency by 3% for a 670 MW plant.

Thereafter, the syngas was heated to 245°C and compressed to 24.4 bar prior to the FT Reactor (Liu *et al.* (2011)). The reactor used was a slurry phased unit with an iron catalyst. Since the refinery section was not modeled in Aspen, the conversion of syngas to fuels (C9-C15) in the FT reactor was specified at 53% to account for product mitigation in the refinery section. Significant reaction heat from the reactor is captured for steam production purposes.

After FT synthesis, the effluent is cooled to facilitate water removal. The production line would then enter the distillation-refinery sub-process (see Section 2.1.4) that is not modeled in detail in Aspen Plus. One of effects of this section though, which was the production of gasses in the refinery was hot wired into the model with a simulator block to reflect its effect on the composition of the light gasses in the model (Kreutz *et al.* (2008)). From this gas stream, 60% is split towards the Auto-Thermal Reformer (ATR) and the remainder is sent to the combustor.

The ATR is operated at 1000°C and 29bar using steam (185°C, 29bar) and pure oxygen from the air separation unit. The flow rate of steam is fixed at 65% (mass) of the gas flow rate while the amount of oxygen is set to fix the adiabatic temperature at a specified level. The ATR reforms approximately 97% of methane and 100% of the other light gasses by converting them to syngas (Leibbrandt (2010). These conditions produced better results than the conditions used by Liu *et al.* (2011), wich only reformed 90% of the methane. The split to the combustor is fed with air pre-heated to 300^{o}C that at 20% in excess of stoichiometry. The combustor generates superheated steam at 38 bar. Steam for the refinery, ATR and Rectisol are drawn off. The remaining steam is expanded at 23.6, 2.4 and 0.046 bar in a series of three electricity generating turbines, where the isentropic and mechanical efficiencies are 85% and 98% respectively (Leibbrandt (2010), Kreutz *et al.* (2008)). After all process electricity is withdrawn, the remainder is exported as a saleable product.

2.1.4 First Order Mass Balance Model for the Distillation-Refinery

Though the detailed mass and energy balance for the distillation-refinery section was not simulated in Aspen Plus, it would be deemed necessary to complete a first order mass balance (i.e. MS Excel Spreadsheet) for these processing sections in the context of the current exercise, because

i. The nature of an energy requirement must be established, because the energy can either be in the form of a fuel gas, steam, electricity or water for cooling. These energy requirements could then be appropriated from the primary Aspen Simulation model the GFT process.

ii. The emissions associated with distillation and all the fuel upgrading units could be estimated.

The process description for the distillation-refinery sub-process was outlined in Bechtel_(c) (1997) for FT syn-crude upgrading. Data pertaining to point (i) is available in three sources, namely:

- Bechtel_(c) (1997) – hydro-treating, catalytic cracker, catalytic reforming. It could be assumed that this data included the steam energy requirements for the debutanizer, or that enough excess steam is generated to suffice the debutanizer. BREF (2003) for example, illustrated that energy generated from the catalytic cracker supplied the debutanizer.
- BREF (2003) – distillation.
- Shah and Fullerton (1990) – isomerisation.

Emission data pertaining to point (ii) above is readily available as averaged values for European refineries in the BREF (2003) report for all air, liquid and solid emissions. Liquid emissions from upgrading sections in refineries result primarily from washing to remove nitrogen and sulphur species that were removed from the organic constituents by hydro-treatment. If there are no sulphur and nitrogen present, as is the case with FT syn-crudes (Bechtel_(c) 1997), then washing is not necessary. However, a water effluent would still arise due to the hydration of oxygenates (Schablitzky *et al.* 2011) in syn-crude fractions. Schablitzky *et al.* (2011) reports the typical concentrations of oxygenates that are to be found in FT syn-crudes. The data for this spreadsheet model is shown in APPENDIX 2.1, Table A 1.

2.2 PINCH POINT ANALYSIS

2.2.1 Literature Review

The energy demand of a biofuel process can very well render a process to be economically unviable because of expenditure on energy utilities. This would be the case for a non-energy efficient process, such as the bioethanol producing processes utilizing acid and liquid hot pretreatment methods in Leibbrandt (2010). Thus, heat integration becomes a necessity to reduce energy costs. In an energy efficient process scenario, such as the Gasification FT process, heat integration is still a necessity in order to create further excess energy in a saleable form (electricity).

Pinch Point Analysis to improve the energy efficiency of prospective bioethanol flow-sheets have been carried out in previous studies, such as Dias *et al.* (2010) and Grisales *et al.* (2005). Both these studies were conducted to minimize the heating and cooling utilities of the plant to achieve savings on energy costs. In the case of Grisales *et al.* (2005), a practical saving of 23.6% was achieved on heating, while 38.7% was achieved on cooling. Dias *et al.* (2010), achieved reductions of 30.8% and 33.7% on high and low pressure steam respectively, while the cold utilities were not reported.

2.2.2 Overview of the Pinch Point Procedure

Dias *et al.* (2010) and Grisales *et al.* (2005) followed the procedures outlined by Kemp (2007). This procedure is a systematic set of heuristics that uses the raw thermal data from a mass and energy balance simulation for the heat integration.

The pinch point analysis begins with the construction of the composite diagrams, which represents the stream thermal data in plots, that profiles the heat availability and demands of the process across the entire range of temperatures (Kemp (2007)). Thus, the amount of overlap between the heat availability profile (i.e. the hot composite curve) and the heat demand profile (i.e. the cold composite curve) presents the amount of heat available for transfer (Peters and Timmerhaus (1997)). Furthermore, the exceeding of each curve beyond the other represents the respective utility requirement (see model graph, Figure 4). These deductions are only ascertained once the minimum distance between the two composite curves is at the specified minimum temperature, i.e. the DTmin.

The second inference from a combined composite diagram (after accounting for the DTmin) is the capital consequences of heat integration, the capital consequences being the cost associated with the purchasing and installation of the heat transfer units required for heat integration (see model graph, Figure 4). This is an “inverse inference” from the area enclosed by the two composite curves and utility segments. Thus, as the DTmin increases, the capital demands decrease while the utility demands increase (vice versa for decreasing DTmin). If heating utilities increases, then the energy efficiency would decrease because the energy input to the plant increases.

Thus, the selection of a DTmin is crucial for the satisfaction of the optimum between the capital and utility demands. The target for DTmin can be achieved through a capital-utility trade-off analysis using a range of values for DTmin as inputs to find an optimum. March (1998) recommended the ranges for DTmin that should be used for certain industries and/or substance usage. These ranges, through experience, take into account the fouling and heat transfer potentials. Based on those recommendations then, a suitable range for DTmin in a Bio-ethanol process would be 15-50K. Thus, the values for DTmin used in the study of Grisales *et al.* (2005), which were 5 and 10K, were below the recommended ranges. This implies that the potential savings indicated in that study was overestimated. Dias *et al.* (2010), however, did not carry out a DTmin comparison, but a fixed DTmin value of 15K was used. It is noteworthy that the DTmin values for bioethanol scenarios are generally minimized because the consequences of utility costs far exceed that of capital costs.

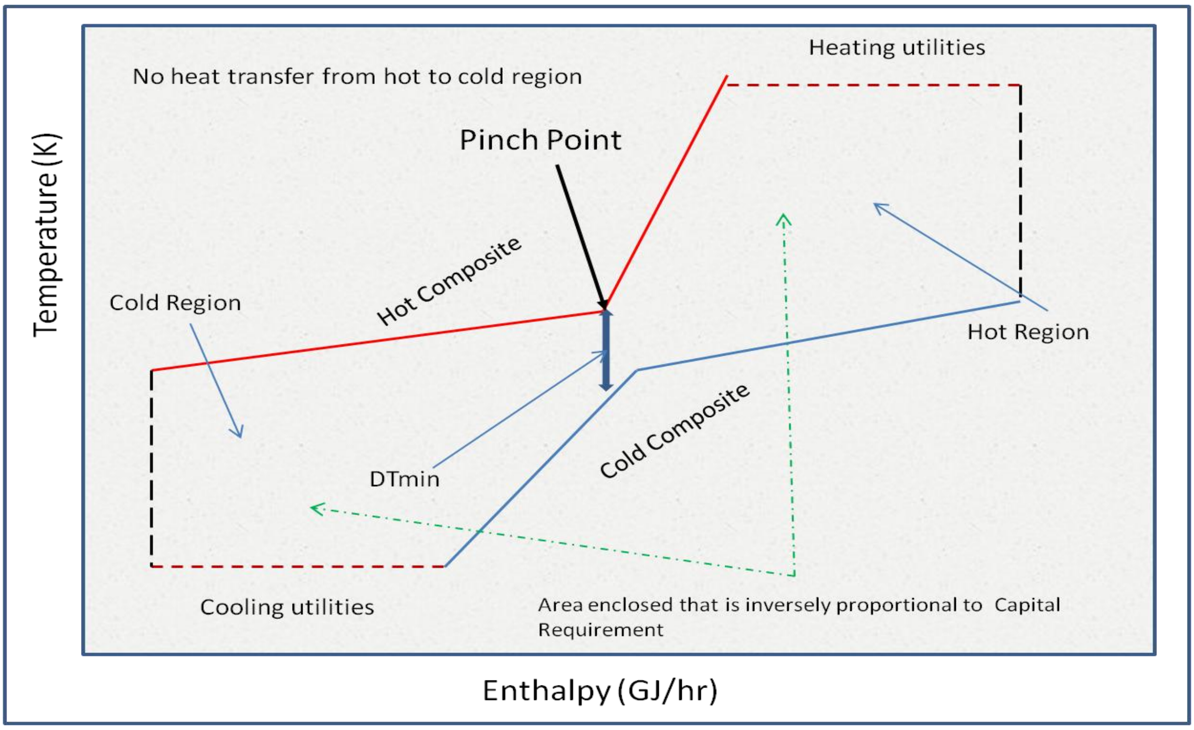


Figure 4: Model Graph to Demonstrate Pinch Point Principles

Pricing data for capital and energy requirements is needed for the capital-utility trade-off analysis. Capital pricing must be representative of the complete costs of heat-exchange equipment while energy pricing must be reflective of the utilities needed. Al-Riyami *et al.* (2001) provided an extensive range of heat exchanger and utility costs. While costs of heating utilities are easy to attain (e.g. coal/natural gas), cooling costs have to be representative of various cooling scenarios, since cooling could be achieved through fin-fanning, cooling towers, sea water coolers etc. These scenarios were all considered by Al-Riyami *et al.* (2001) when establishing the cold utility cost. In order to carry out the analysis, the capital cost is annualized (Kemp (2007)). This is done by inflating the proposed investment over the period of concern and then converting it into a yearly amount.

Once a DTmin has been established along with its utility targets, the next step would be to construct a Heat Exchanger Network (HEN). The procedure for constructing the network is outlined in Kemp (2007) and also in Peters and Timmerhaus (1997). The basic objectives of this procedure are to design a network to achieve the minimum utility target while minimizing the amount of unit exchangers that are used. These must be achieved without violating the following rules:

1. Heat cannot be transferred across the pinch (see model graph, Figure 4) .
2. Hot utilities cannot be used below the pinch and vice versa for cold utilities (see model graph, Figure 4) .
3. The approach temperatures of streams entering a unit cannot be less than DTmin.

4. Two streams cannot have more than one unit of heat exchange.

The absolute minimum number of units is defined by the sum of the process streams (including the virtual streams representing hot and cold utilities) less one. This minimum is not always achievable though, since the pinch point divides the entire heat exchange region into two thermally independent regions (Perrins 1994).

2.3 AUXILLARY PLANT DESIGN

2.3.1 Cooling Tower

Cooling towers are to be used to cool the hot effluent to 35^{o}C, which is necessary since environmental regulations (Water-Act-1956 1984) stipulates that discharge into water bodies cannot exceed this standard. If the effluent is sent for water treatment, this standard would still necessary because the biological processes involved in the purification of water would cease to function at high temperatures.

The key mechanism in a cooling tower is evaporative cooling. Therefore, cooling towers could be modelled as flash vessels in Aspen Plus, where the effluent temperature of 35^{o}C is specified, and the evaporative loss is also specified, which was determined by the following formula (Perry and Green 1997):

$$W_e = 0.00085 \times W_c \times (T_1 - T_2)$$

Where We - evaporative loss.

Wc- circulating-water flow, gal/min at tower inlet

T1 – T2 - inlet-water temperature minus outlet-water temperature, °F

The resulting heat duty is scaled by the Coefficient of Performance to determine the electrical requirement for the pumps and fans that operate the cooling tower. The COP of a cooling tower will have seasonal variations, and thus an average value is to be ascertained for modelling purposes (Bergstein 2009). Bergstein (2009) used a conservative annual average of 7.

2.3.2 Water Treatment Plant

It could be assumed that the water treatment plants connected to all process scenarios follows the standard flow-sheet described in Cyrus (1971). In the context of this study however, these processes would not be modelled in any detail whatsoever, but their energy usage is needed for consideration in terms of the overall energy balance. Smith (1973) provided data that can be modelled to estimate the overall energy needs of the water treatment plant.

2.4 METHODOLOGY

2.4.1 Mass and Energy Balances

Aspen Plus simulation models for the mass and energy balances of the current processes were developed by Leibbrandt (2010). For each advancement, or desired change, a new scenario would be created by modifying the processes of Leibbrandt (2010) to reflect the change in the mass and energy balance to determine the process needs, outputs and inputs and utilities. Since there were no notable advances on the Gasification-FT model, this was left as is.

The modifications would be based on experimental data from literature. The data that would be implemented includes operating conditions, chemical use and dosages, and the conversions obtained.

Hence, the modifications:

1. Steam Explosion catalysed by sulphur dioxide rather than sulphuric acid. Data from Carrasco *et al.* (2009) would be used for this advancement.
2. Simultaneous Saccharification and Co-Fermentation. Experimental data from Rudolf *et al.* (2007) would be used. The data would extend into reconfiguring the yeast cell propagation scheme.

For the modeling purposes in Aspen Plus, the ELECNRTL thermodynamic method will be applied wherever electrolyte solutions occur (Leibbrandt (2010)). For the distillation and energy supply sections, the NRTL method will be used since it describes phase equilibrium of strongly non-ideal solutions very well.

Since the FT model does not involve any electrolytes, the NRTL method will be the underlying thermodynamic model throughout. The mass balance for the distillation-refinery section will be done according to the literature review in section 2.1.4.

2.4.2 Pinch Point Methodology

2.4.2.1 Process Stream and Thermodynamic Data

All process stream data and subsequent thermodynamic data were gathered from the Aspen-Plus Simulation and the thermodynamic database in Aspen Plus. All heat demanding streams were considered for the analysis while only the streams with meaningful excess heats were considered. Thus, heat excesses that were around than 0-2% of the total heat availability were not considered if the temperatures were significantly lower than the pinch temperature (Dias *et al.* (2010)). This minimum temperature is found iteratively by inspection of a rough composite diagram.

2.4.2.2 **Composite Curve Construction and Pinch Point Location**

The hot and cold composites were constructed and manipulated to find the pinch temperature. The manipulation was to offset the cold composite curve so that it only touched the hot composite curve at a unique value. This offset could be ascertained using packages such as Aspen Pinch (Grisales *et al.* 2005) or by the tabulating method described by Linnhoff *et al*. (1982) (as cited by Dias *et al.* (2010)). The manual for Pinch Point Analysis by Linnhof *et al*. (1982) is the earlier edition of Kemp (2007).

2.4.2.3 **DTmin Optimization for Capital-Utility Trade-off**

The table generated by the "Linhoff method" was modified to account for the values of DTmin within the recommended ranges. The modification procedure is also detailed in Kemp (2007). With each new DTmin that was considered, the utility requirements were calculated from the overlaps of the composite curve. In order to have calculated the heat exchange area, the area enclosed by the composite curves and the utility segments was discretised. Each discrete section was treated as a separate heat exchanger for which the area was to be calculated, and after which, the areas were summed up. The net effect of the cost-saving of utility reduction and annualized capital cost was determined using pricing data extracted from Al-Riyami *et al.* (2001) at each DTmin instance. Thus, an optimized DTmin was to be found in this manner.

2.4.2.4 **Heat Exchanger Network Construction**

The heuristics detailed in Kemp (2007) and Peters and Timmerhaus (1997) was used to construct the Heat Exchanger Network Construction (HEN) without violating any of the prescribed laws. As mentioned before, the objectives were to design a network that achieved the minimum utility target while minimizing the amount of unit exchangers that are used.

A concerted effort to reach the absolute minimum number of exchanger units for the entire region heat transfer region was not considered as such lengths were not considered to be part of this study. Moreover, as the pinch point splits the entire heat transfer region into two thermally independent regions (Perrins 1994), an absolute minimum is not always achievable. It was imperative though, that the minimum number of units be attained on either side of the pinch.

2.4.3 Comparison of Processes

Processes will be compared in terms of the energy efficiency. The two types of energy efficiencies considered are:

1. Liquid Fuel (Energy) Efficiency ($\eta_{liquid\,fuel}$)

$$\eta_{liquid\,fuel} = \frac{HHV_{fuel\,product}}{HHV_{input} - HHV_{by-product} - \dfrac{Elect.Export}{\eta_{45\%}}}$$

Equation 1

The liquid energy efficiency measured the ability of a process to convert the thermal energy in the biomass feedstock ($HHV_{feedstock}$) into the energy present in the liquid fuel products. This definition for energy efficiency adjusts the energy inputs by subtracting the thermal energy ($HHV_{by\text{-}product}$) of bi-products from the thermal energy in the biomass feedstock (Hamelinck *et al.* 2005). Since this energy ratio uses a thermal basis, the electricity that is exported is converted to the hypothetical amount of thermal energy that would be used to generate that electricity. The assumed electrical efficiency (η_{elec}) is 45% (Hamelinck *et al.* 2005).

2. Overall Energy Efficiency ($\eta_{overall}$)

$$\eta_{overall} = \frac{HHV_{fuel} + HHV_{by-product} + Elect.Exdport}{HHV_{input} + Fossil\_Energy}$$

Equation 2

The overall energy efficiency describes the performance of the process of the total energetic output, relative to an energy input that is adjusted to account for any additional energy source (Leibbrandt 2010).

2.5 SUMMARY OF FLOW DIAGRAMS AND MASS BALANCES

2.5.1 Biological Scenarios

2.5.1.1 Generic Block Flow Diagram

The overall block flow diagram for the biological scenarios is given below. Refer to Appendix 1.1 for the detailed process flow diagrams of each hierarchy (area).

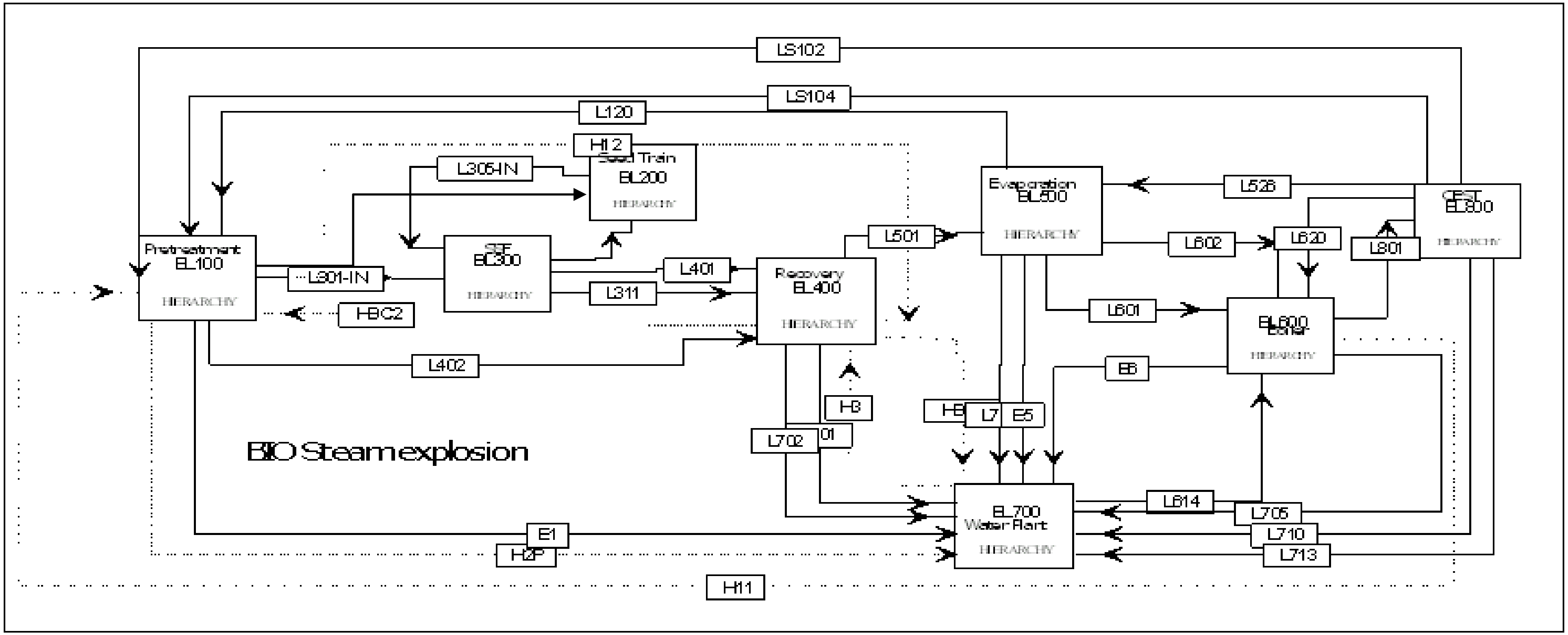
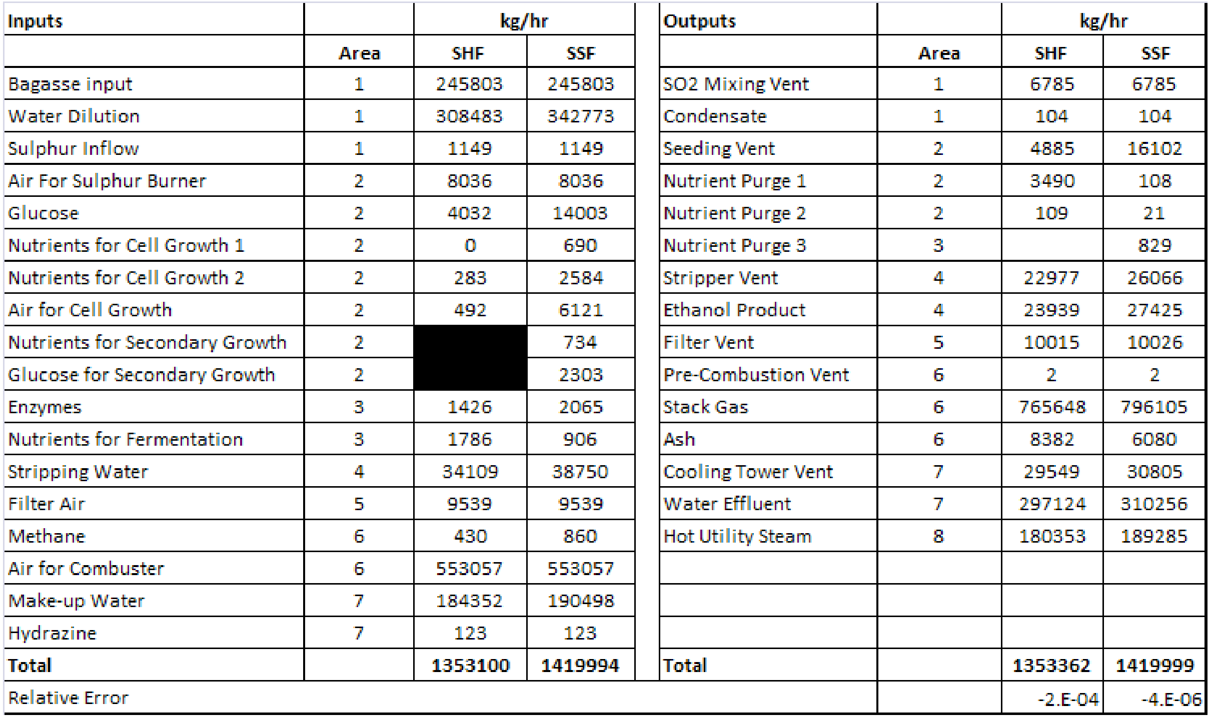


Figure 5: Generic BFD for Biological Scenarios

2.5.1.2 Summary of Mass Balance – Inflows and Outflows

THE OVERALL MASS BALANCES FOR THE BIOLOGICAL SCENARIOS ARE GIVEN BELOW. REFER TO Appendix 1.1 for the detailed stream data of each hierarchy (area).

Table 2: Overall Mass Balance of Biological Scenarios



Inputs		**kg/hr**		**Outputs**		**kg/hr**	
	Area	**SHF**	**SSF**		**Area**	**SHF**	**SSF**
Bagasse input	1	245803	245803	SO2 Mixing Vent	1	6785	6785
Water Dilution	1	308483	342773	Condensate	1	104	104
Sulphur Inflow	1	1149	1149	Seeding Vent	2	4885	16102
Air For Sulphur Burner	2	8036	8036	Nutrient Purge 1	2	3490	108
Glucose	2	4032	14003	Nutrient Purge 2	2	109	21
Nutrients for Cell Growth 1	2	0	690	Nutrient Purge 3	3		829
Nutrients for Cell Growth 2	2	283	2584	Stripper Vent	4	22977	26066
Air for Cell Growth	2	492	6121	Ethanol Product	4	23939	27425
Nutrients for Secondary Growth	2		734	Filter Vent	5	10015	10026
Glucose for Secondary Growth	2		2303	Pre-Combustion Vent	6	2	2
Enzymes	3	1426	2065	Stack Gas	6	765648	796105
Nutrients for Fermentation	3	1786	906	Ash	6	8382	6080
Stripping Water	4	34109	38750	Cooling Tower Vent	7	29549	30805
Filter Air	5	9539	9539	Water Effluent	7	297124	310256
Methane	6	430	860	Hot Utility Steam	8	180353	189285
Air for Combuster	6	553057	553057				
Make-up Water	7	184352	190498				
Hydrazine	7	123	123				
Total		**1353100**	**1419994**	**Total**		**1353362**	**1419999**
Relative Error						-2.E-04	-4.E-06

2.5.2 GFT Scenario

2.5.2.1 Process Flow Diagram

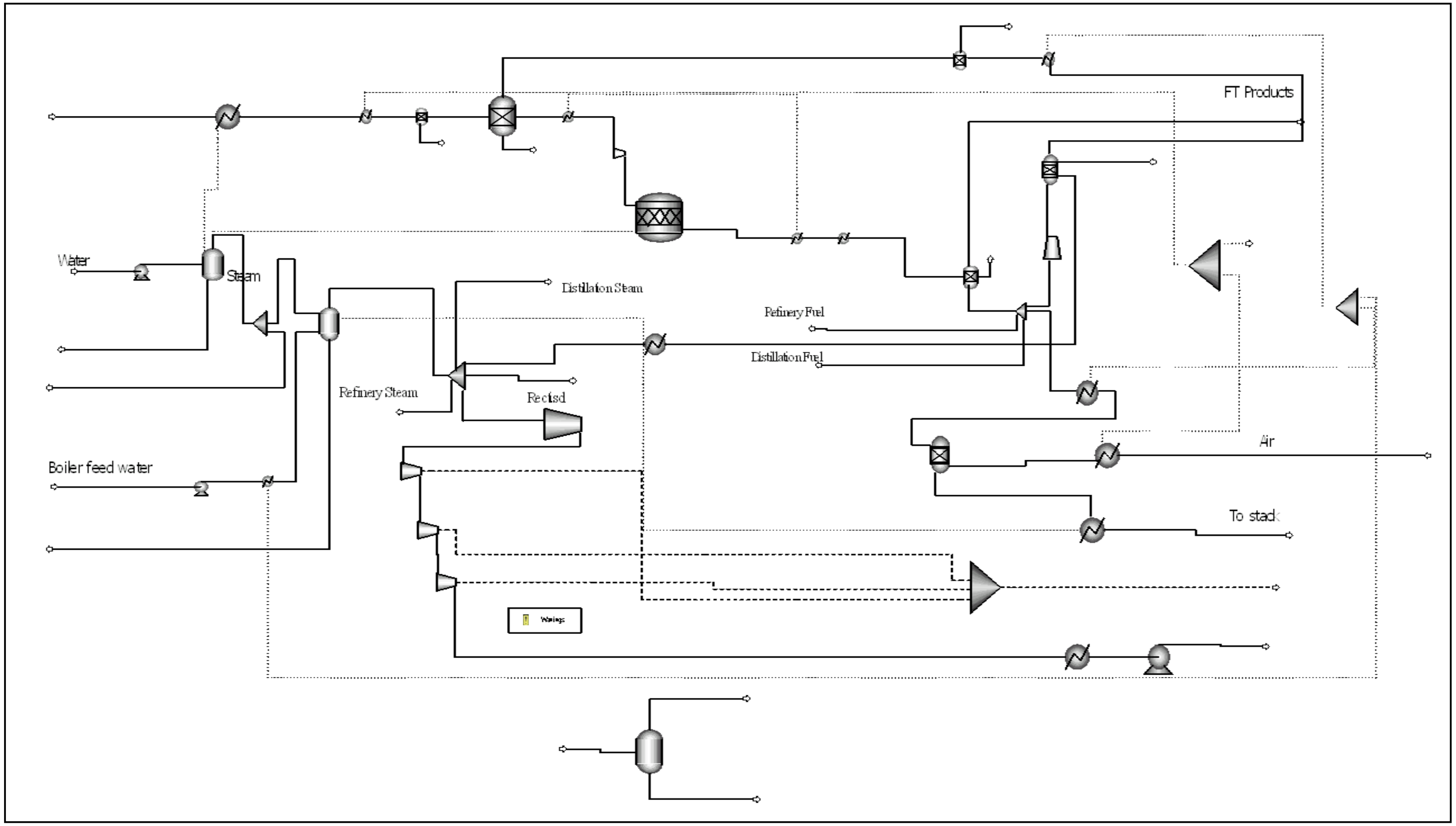
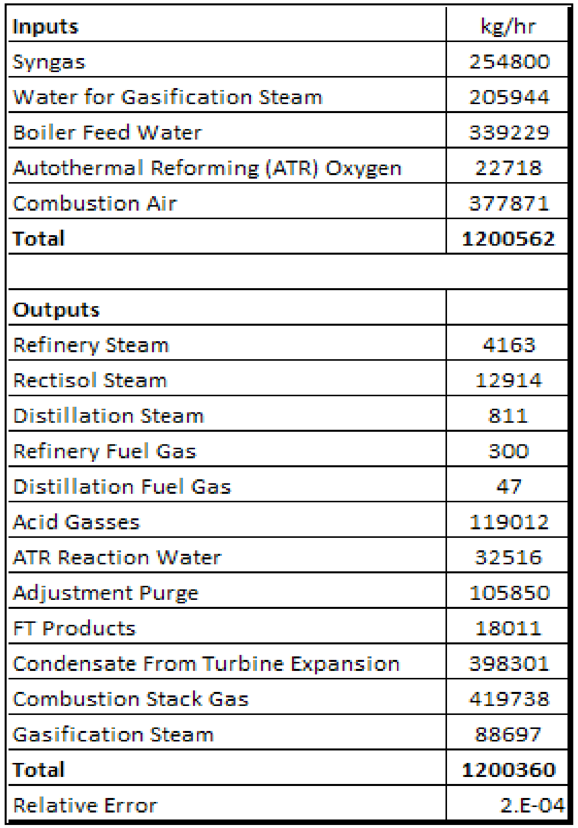


Figure 6: PFD for GFT Scenario

2.5.2.2 Summary of Mass Balance – Inflows and Outflows

The overall mass balance for the GFT Scenario is given below. Refer to Appendix 1.2 for the detailed stream data.

Table 3: Overall Mass Balance for GFT Scenario



Inputs	kg/hr
Syngas	254800
Water for Gasification Steam	205944
Boiler Feed Water	339229
Autothermal Reforming (ATR) Oxygen	22718
Combustion Air	377871
Total	**1200562**
Outputs	
Refinery Steam	4163
Rectisol Steam	12914
Distillation Steam	811
Refinery Fuel Gas	300
Distillation Fuel Gas	47
Acid Gasses	119012
ATR Reaction Water	32516
Adjustment Purge	105850
FT Products	18011
Condensate From Turbine Expansion	398301
Combustion Stack Gas	419738
Gasification Steam	88697
Total	**1200360**
Relative Error	2.E-04

2.6 RESULTS AND DISCUSSION

*For all sections, the results for the SSF and SHF processes will be presented together to allow easy comparisons between these two processes that are fundamentally the same.

2.6.1 Pinch Point Analysis of Biological Processes

2.6.1.1 **Stream Selection**

The streams considered for pinch point are shown in Table 4. Stream locations are indicated in the PFD diagrams of the relevant areas (shown in brackets) in APPENDIX 1 and the thermal data is given in detail in APPENDIX 2.3, Table A 5.

Table 4: Stream Selection for Biological Models

		SHF Configuration				SSF Configuration			
H	**Hot Streams (Availability)**	**Mass Flow Rate**	**Duty**	**Initial Temperature**	**Target Temperature**	**Mass Flow Rate**	**Duty**	**Initial Temperature**	**Target Temperature**
		t/hr	**MW**	K	K	**t/hr**	**MW**	K	K
1	Pre-Steamer Vent (Area100)	9.45	-6.62	308.15	420.00	9.49	-6.65	308.15	420.00
2	SSF/Fermt. Feed (Area300)	630.50	-18.70	304.15	338.15	687.87	-5.01	304.15	312.00
3	Rectifier Condenser (Area400)	73.14	-19.40	359.00	359.00	79.62	-21.18	359.00	359.00
4	Beer Column Condenser (Area400)	156.32	-69.30	375.00	375.00	169.07	-75.42	375.00	375.00
5	Area 600 Bleed	25.61	-4.68	308.15	464.70	26.87	-4.92	308.15	465.00
6	Area 700 Bleeder 1	24.40	-16.97	308.15	410.15	23.48	-16.29	308.15	406.74
7	Area 700 Bleeder 2	218.32	-43.93	308.15	373.15	230.13	-46.31	308.15	373.15
8	Etoh-Product (Area400)	18.98	-5.11	308.15	357.50	21.74	-5.85	308.15	357.50
C	**Cold Streams (Burden)**	**Mass Flow Rate**	**Duty**	**Initial Temperature**	**Target Temperature**	**Mass Flow Rate**	**Duty**	**Initial Temperature**	**Target Temperature**
		t/hr	**MW**	**K**	**K**	**t/hr**	**MW**	**K**	**K**
1	Bagasse feed (Area100)	245.80	18.35	298.15	400.33	245.80	18.35	298.15	400.33
2	1st CO2 Flash (Area400)	609.42	29.29	319.15	362.41	664.14	32.47	319.15	363.16
3	Beer Column Feed (Area400)	36.68	2.54	295.00	354.00	41.74	2.89	295.00	354.00
4	Boiler Water Make-Up(Area700)	184.35	13.42	298.15	373.15	160.10	0.01	298.15	373.15
5	Beer Column Reboiler (Area400)	149.80	92.23	388.35	388.35	164.16	101.09	388.35	388.35
6	Rectifier Reboiler (Area400)	18.28	11.25	387.47	387.47	19.17	11.79	387.47	387.47

The heat required to heat the Bagasse Feed is heat required to maintain the Pre-Steaming unit at its design temperature. Thus, an isothermal heat load at an otherwise high temperature has been transferred to a lower temperature state. This relaxes the quality of the heat demand, and would result in the recovery of heat from hot streams at lower temperatures, which would otherwise need removal through a cooling utility. The same rationale was applied for both the water make-up stream to relax the intensity of heat demand of the De-aerator, and the CO_2 Flash unit feed.

Regarding the hot streams, heat recovery was sourced primarily from hot liquid effluents and steam vents, such as that of the De-aerator and Pre-Steaming units. Otherwise, heat availability arises from condenser duties and the ethanol product. Heat availability was also placed before the fermentation tank to lessen the duty on the refrigeration system that acts through a submerged coil in the vessel, to remove reaction and other sensible heat.

2.6.1.2 **Composite Curves and Pinch Location**

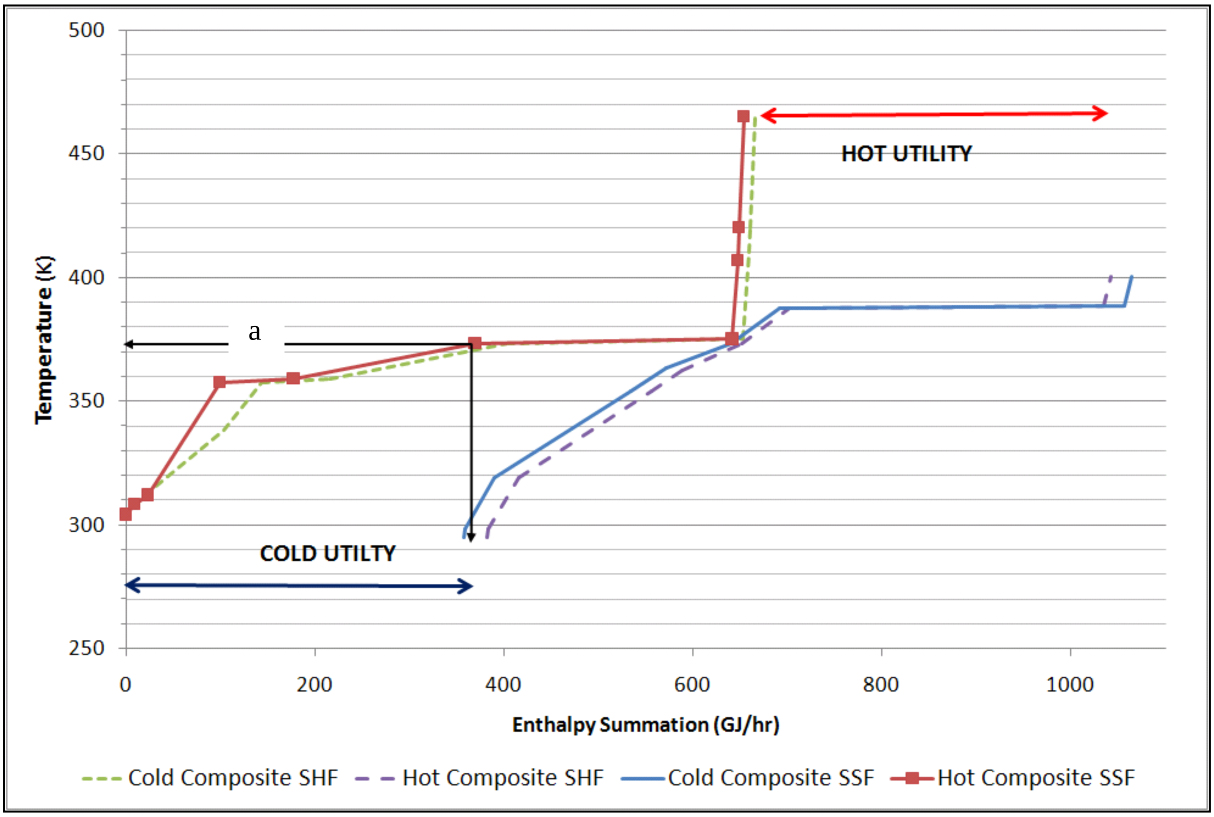


Figure 7: Composite diagram for Biological Processes

The composite curves for the SSF and SHF processes shown in Figure 7 indicate that the pinch point temperature lies at 375K for both the SSF and SHF. This temperature level was primarily a result from the large amount of heat arising from the condenser of the beer column. To some extent, the heat of condensation that was available in the steam that arose from vents at atmospheric pressure also forced the pinch point to occur at this temperature. The duty of this condenser is thus recoverable amongst a number of cold steams below the pinch.

The composite curve diagram (Figure 7) shows that minimum effective temperature of a hot stream must be greater than 370K, for its heat to be captured by a cold stream below the pinch. (This concept was indicated by the line called “a” in Figure 7, which connected the lower end of the cold composite curve to the hot composite, and then to the corresponding temperature, for the SHF scenario). Thus, majority of the heat availability below the pinch point would consequently be dispensed to the cooling utilities. Even a large heat source, such as the condenser duty of the Rectifier, was rendered irrecoverable.

As soon as the DTmin was increased, the value of the minimum effective temperature for the hot steams increased and the amount of overlap between the composite curves decrease. This shows how the level of heat integration diminishes as the DTmin increased. At the same instance, the total area enclosed by the curves and utility lines increased, this implied that the requirement of heat transfer area also diminished. These effects are shown graphically for the SSF Scenario in Figure 8 as an example. A similar figure for the SHF scenario is in the APPENDIX2.3, Figure A1.

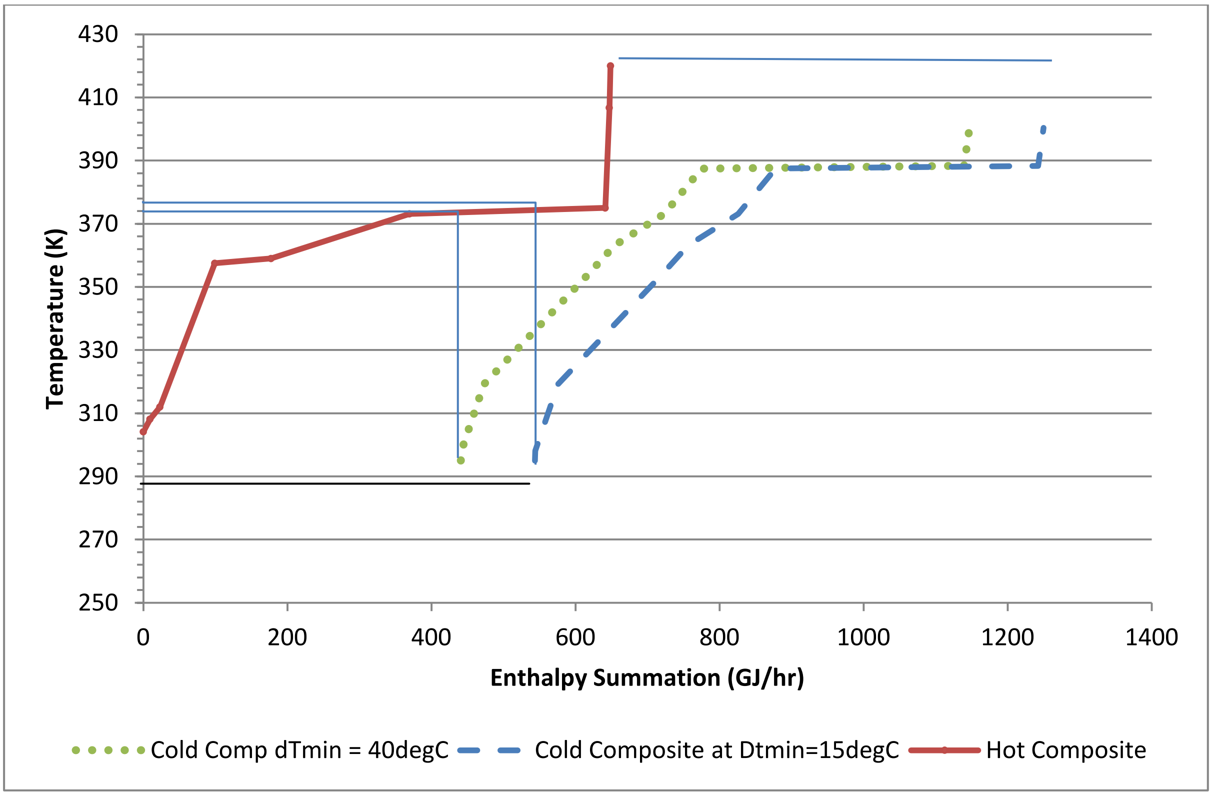


Figure 8: Effect of DTmin on SSF Composite Diagram

Above the pinch, heat is available at very high temperatures, but unfortunately, in small amounts that do very little to satisfy the heat demands above the pinch. There is no heat available for the distillation reboilers, which is indicated by the flattened sections of the cold composite curve, that account for >50% of the total heat demand.

A possible solution to increase the quality of the heat availability would be to retrofit the vents of the Pre-Steaming and Deaerator units with high pressure steam pipes that will allow for installation of heat exchanges at high pressures, instead of an immediate throttle valve that reduces the steam pressure to 1atm. This proposed modification will enable the heat of condensation to be captured at temperatures as high as 400-410K. A cost-benefit analysis should be carried out though to assess the feasibility of such an action.

2.6.1.3 **DTmin Optimisation**

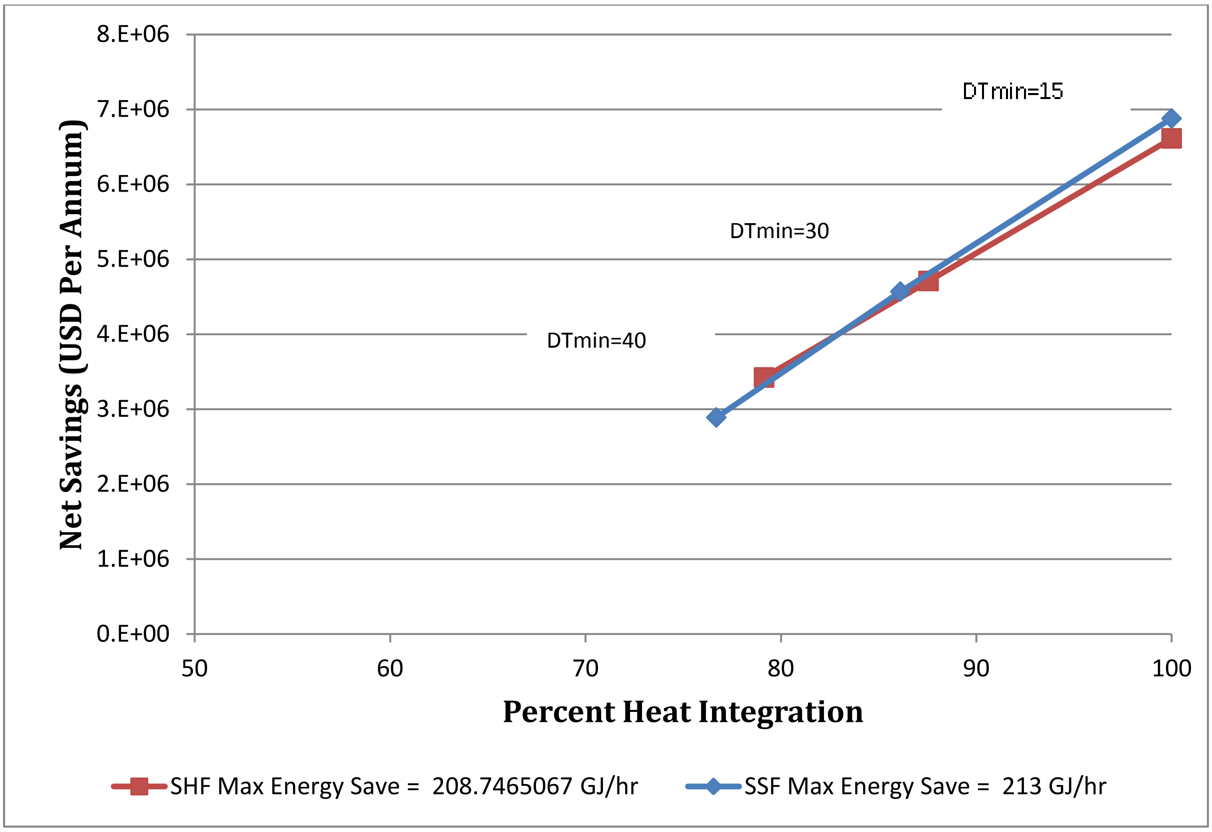


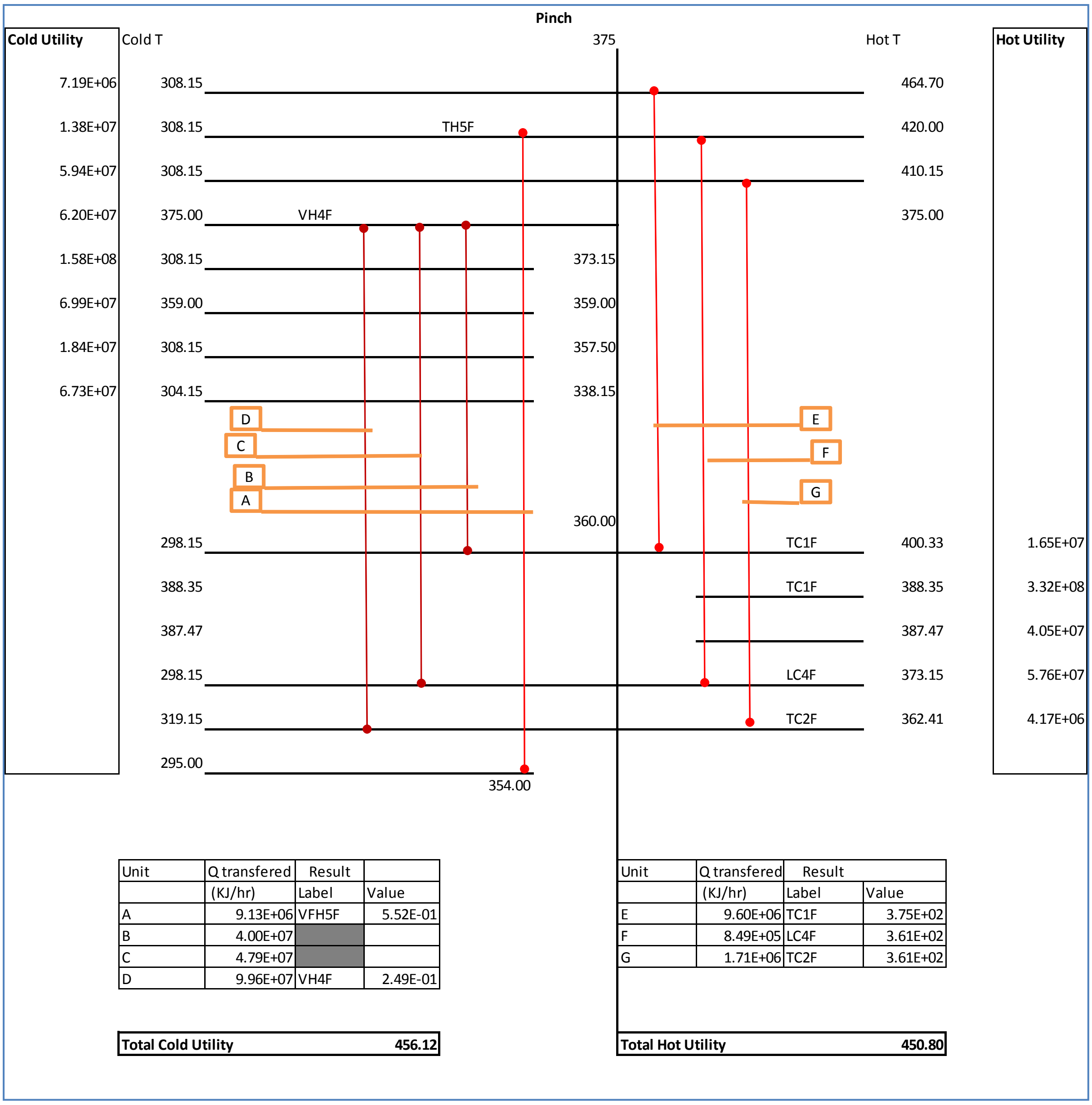
Figure 9: DTmin Optimisation i.t.o Heat Integration

The results of DTmin optimisation for the biological processes are shown in Figure 9. The "net saving" parameter on the y-axis in Figure 9 was calculated as the difference between the savings incurred from lower operating expenditure on utilities and the annualized capital costs. The level of heat integration is directly related to the DTmin, because 100% heat integration meant that the minimum approach temperature between hot and cold streams was 15K, where heat recovery is at a maximum, but the heat transfer area is also a maximum. Correspondingly, the mid-points on the curves (+/- 85-90% heat integration) referred to a DTmin of 30K and the lower point refers to a DTmin of 40K. Thus, the SSF and SHF results show that heat integration should be maximised, else said, the DTmin be set at 15K. The trends shown in Figure 9 indicates that the decrease in capital requirements is insignificant compared to the increase in utility costs that arose when heat integration was compromised.

2.6.1.4 **Heat Exchanger Network Design**

The heat exchange network for the SHF process is shown in Figure 10. Refer to APPENDIX 2.3, Figure A2 for the network of the SSF process. The design of the heat exchange networks is shown to have met the utility requirements at the specified DTmin. The design of the networks also confirms the deductions of the composite diagrams, which were:

- Heat availability from a hot stream was only valuable at temperatures greater than 370K.
- The heat availability above the pinch was minute compared to the demand and did not full any of the significant requirements.



Unit	Q transfered (KJ/hr)	Result Label	 Value
A	9.13E+06	VFH5F	5.52E-01
B	4.00E+07		
C	4.79E+07		
D	9.96E+07	VH4F	2.49E-01

Total Cold Utility	456.12

Unit	Q transfered (KJ/hr)	Result Label	 Value
E	9.60E+06	TC1F	3.75E+02
F	8.49E+05	LC4F	3.61E+02
G	1.71E+06	TC2F	3.61E+02

Total Hot Utility	450.80

Figure 10: HEN of the SHF Process

The number of exchange units, though minimised on either side of the pinch point, could not reach the hypothesised number of 15units, because the potentials for any once off stream annihilation were nullified due to the following reasons:

- the majority of target temperatures of the hot streams could not approach the initial temperatures of the cold streams greater than above the DTmin.
- The heat capacity potentials of the cold streams were much greater than that of the hot streams, which would have lead to an eventual pinch violation.

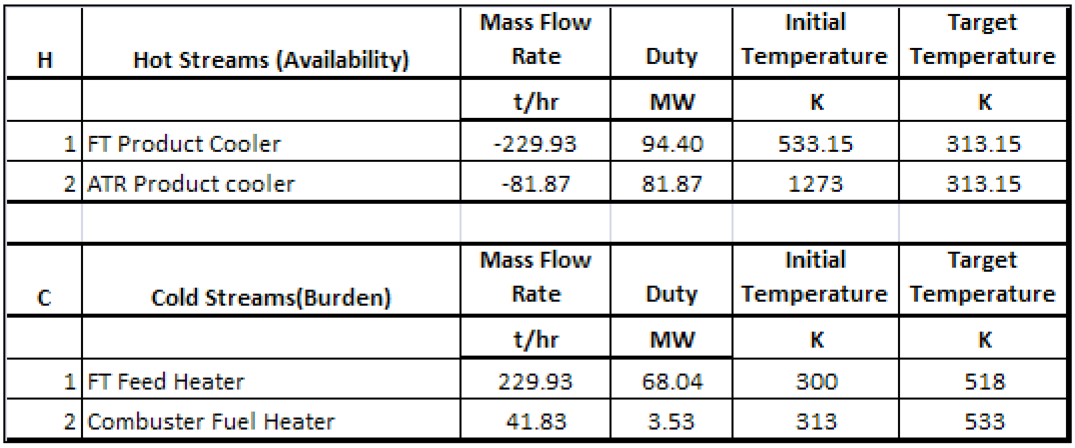
As discussed in Perrins (1994), the minimum number of exchangers that is hypothesised, is hardly obtainable, since the pinch point creates two thermally independent regions.

2.6.2 GFT Process Pinch Point Results

2.6.2.1 **Stream Selection**

The streams selected for pinch point analysis are shown in Table 5. These were considered for the gasification and FT sections of the plant. The full details of the stream data is given in APPENDIX 2.3, Table A 5.

Table 5: Stream Selection for GFT process



H	Hot Streams (Availability)	Mass Flow Rate	Duty	Initial Temperature	Target Temperature
		t/hr	MW	K	K
1	FT Product Cooler	-229.93	94.40	533.15	313.15
2	ATR Product cooler	-81.87	81.87	1273	313.15
C	**Cold Streams(Burden)**	**Mass Flow Rate**	**Duty**	**Initial Temperature**	**Target Temperature**
		t/hr	MW	K	K
1	FT Feed Heater	229.93	68.04	300	518
2	Combuster Fuel Heater	41.83	3.53	313	533

The cooling duties were required for low temperature separation processes, such as Rectisol acid gas removal, since gas solubility is maximised at low temperatures (i.e Henry's Law), and the separation of water from the FT reactor effluent. Heat duties are required to preheat reaction mixtures to the reaction activation temperatures. Pre-heating the combustor feed maximises the adiabatic temperature of the flue gasses.

Streams in the Refinery section were not included in the pinch point analysis, because the refinery in the model from Leibbrandt (2010), was only modelled on a first order basis and thus, the detailed thermal information required for pinch point analysis was not available.

2.6.2.2 **Composite Curves and Pinch Point**

The composite curves for the SSF and SHF processes are shown in Figure 11. The composite diagram shows that the pinch point exists at 533K. There is an abundance of heat available from the hot streams to satisfy all the demands of the cold streams.

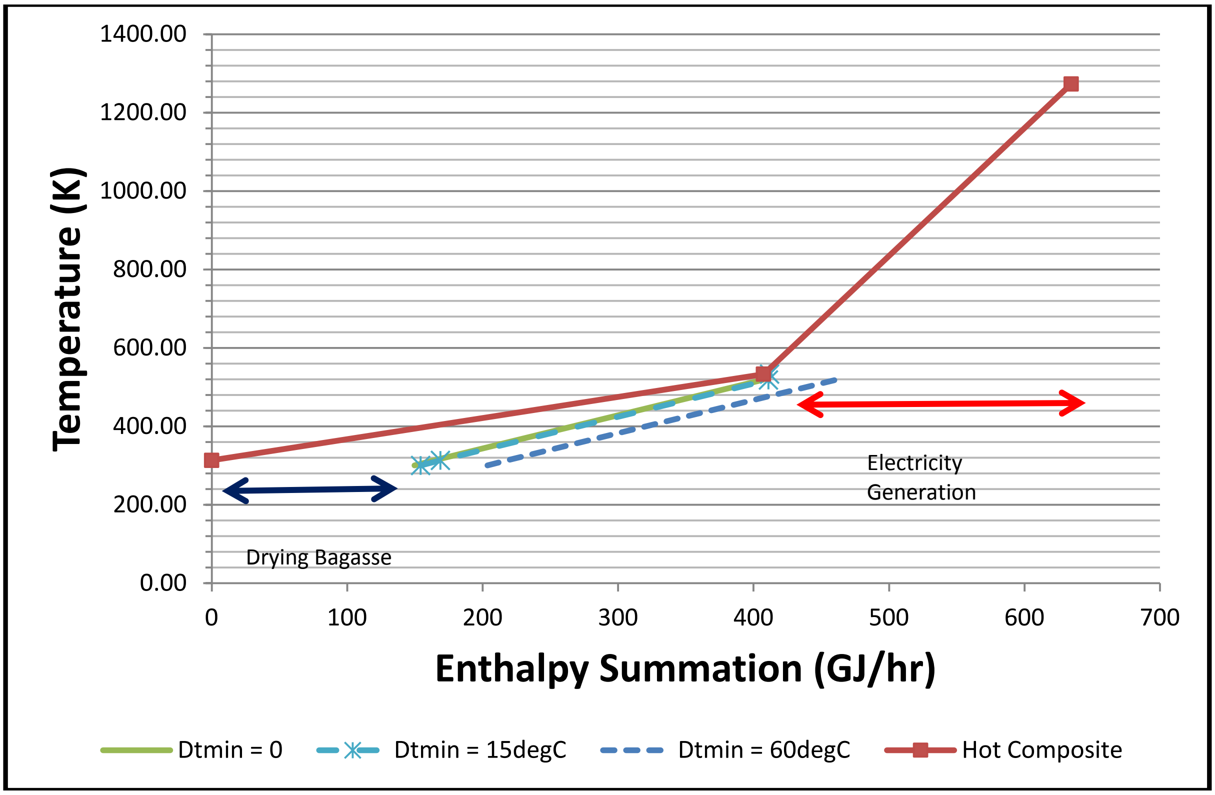


Figure 11: Composite diagram for GFT process

In addition, the region above the pinch point reveals that there is no heat demand. Thus, a surplus of high temperature heat is available for beneficial purposes. This surplus heat would be used to generate extra electricity to boost the sales revenue.

2.6.2.3 DTmin Optimisation

The results of the DTmin optimisation, with respect to the GFT scenario are presented in Figure 12. The values of the DTmin's considered in this process was 15K (100%), 40 (mid-point) and 60K (minimum integration).

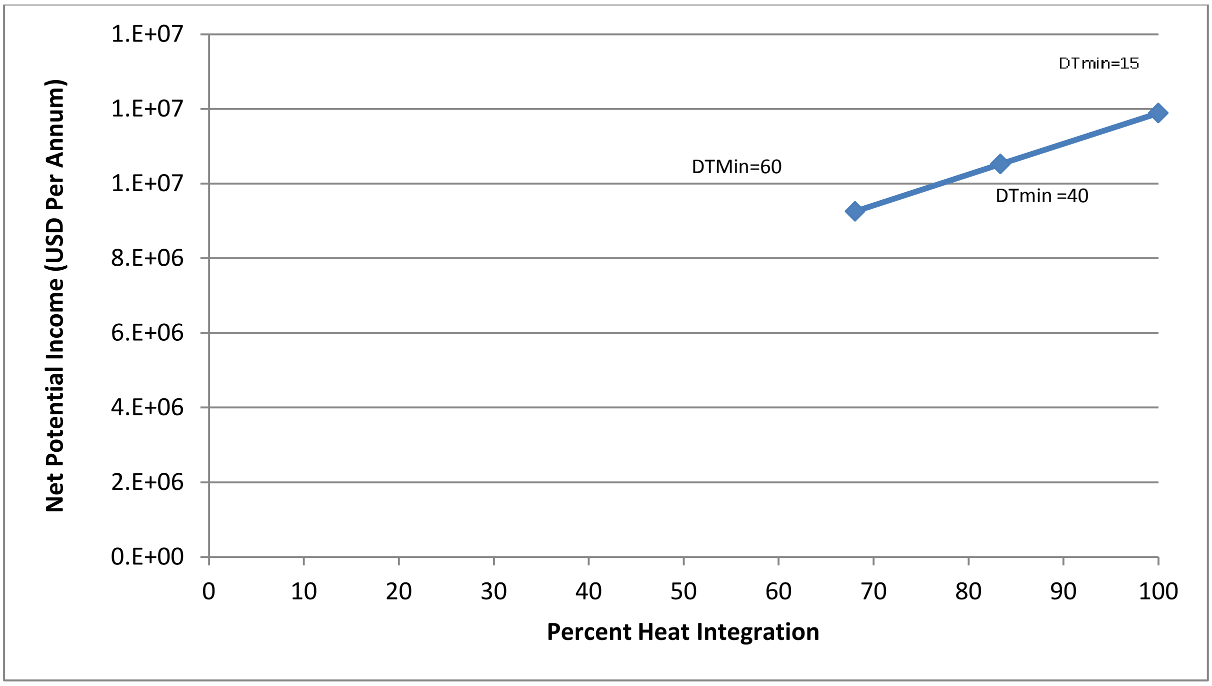


Figure 12: DTmin optimisation for GFT scenario

As was with the biological processes, the maximum heat integration was targeted. Capital costs were also rendered inconsequential compared to the utility costs, as shown by the trend in Figure 12. In this scenario however, the pinch point analysis did not just yield a significant energy saving, but a generation of large amounts of super-heated steam (see Section 2.6.2.2) to be used for electricity generation. Figure 11 indicates that the amount of high-temperature heat available for generating electricity at a DTmin of 15K is 222 GJ/hr and that would amount to 89 GJ/hr after an electrical conversion efficiency of 40% was taken into account. The energy efficiency applied here is reflective of an expansion circuit of a fossil fuel power plant, without the losses associated with the boiler, since the form of energy that is under consideration here is already that of heat. Thus, if a power plant that has an overall efficiency of 32.5% (Kreutz *et al.* (2008)), includes a boiler that has an efficiencies ranging from 75-85%, then the efficiency of the expansion circuit by itself would range from 38% to 43%.

2.6.2.4 Heat exchange Network Design

The minimum number of units achievable for the system is 5, and 5 units were achieved. Unlike the restriction in the biological processes, a hot stream was able to satisfy the entire demand of a cold stream because it had a greater heat capacity. The heat exchange network for this scenario is shown in APPENDIX 2.3, Figure A3.

2.6.3 Process Energy Demands

2.6.3.1 Energy Requirements for Biological Processes

Figure 13 and Figure 14 shows the specific energy requirements, in terms of MJ per kilogram of bagasse feed, for each of the major divisions of the SHF and SSF process scenarios. These results are presented in terms of the electrical (Figure 13) and heat (Figure 14) energy demands in a comparative manner. Refer to APPENDIX 2.1 (Table A 2-4) for the energy requirements for the major processing units.

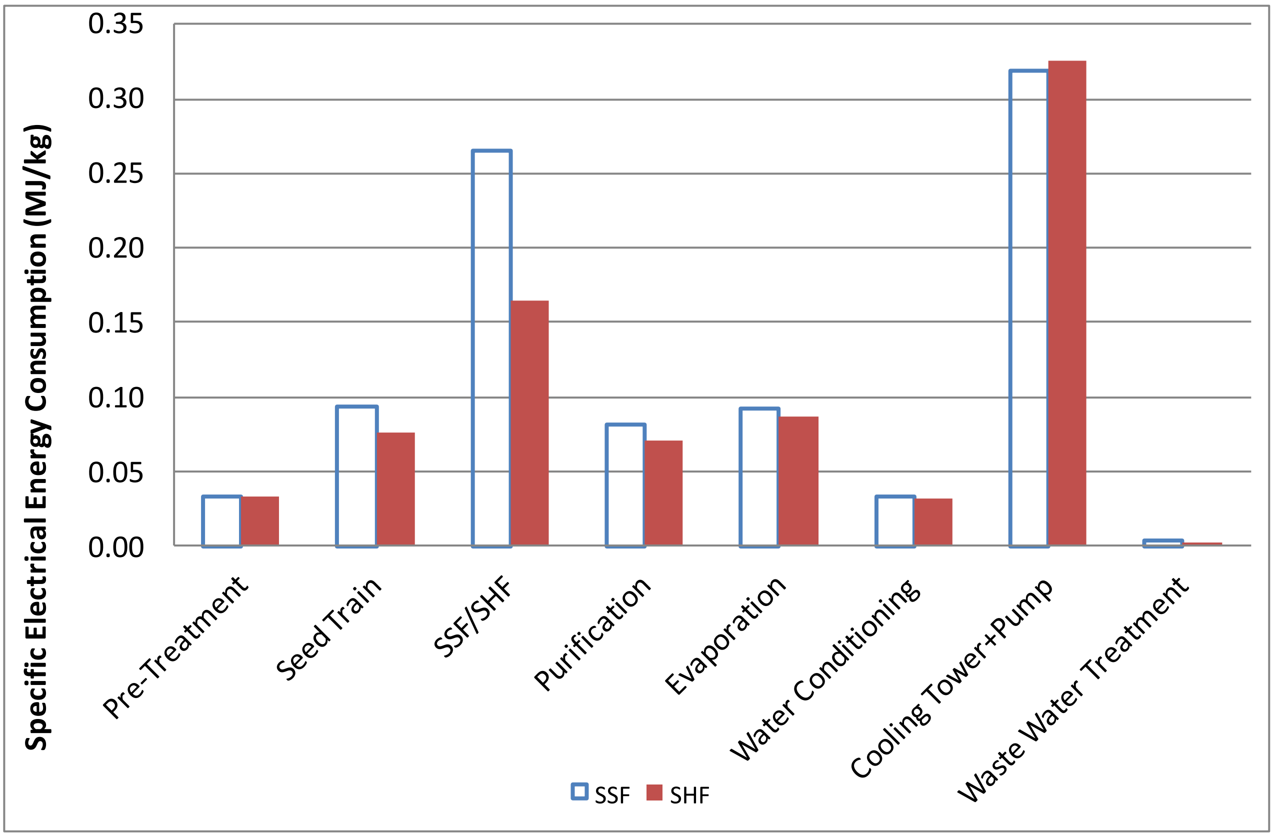


Figure 13: Biological Processes Electrical Demands

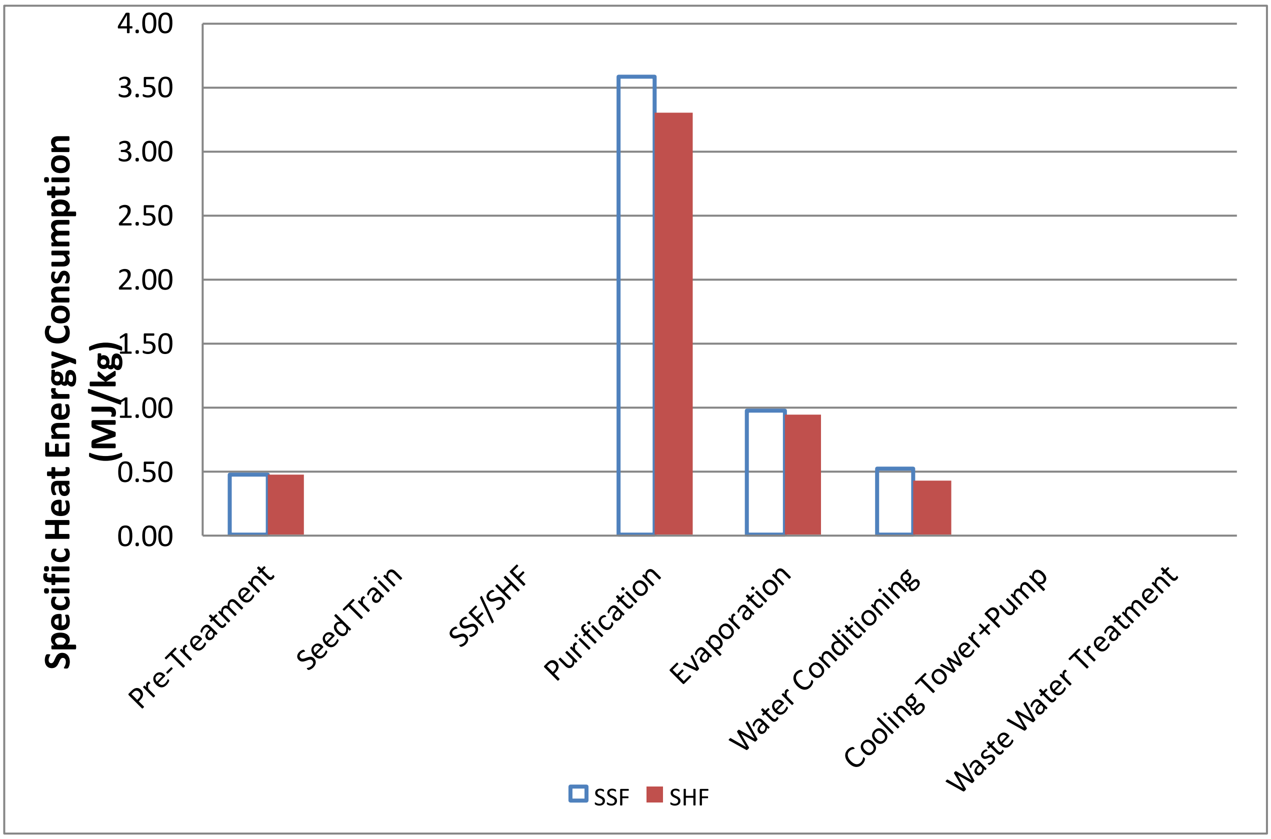


Figure 14: Biological Processes Heat-Thermal Demands

The electrical and heat demands are identical for the pre-treatment step for both processes, as this process remained unaltered. While there were no thermal needs for the Seed Train in either scenario, electricity was needed for the refrigeration requirements of the Seed Trains. The electricity requirement for refrigeration was calculated by factoring the heat duty of the reacting vessel which was determined by Aspen by the COP of the refrigeration system. The cell propagation scheme operated at temperatures that were too low for cooling water, and thus, refrigeration was employed.

For the hydrolysis-fermentation sections, Figure 13 shows the SSF technology demanded more electricity for refrigeration than the SHF technology. This occurred because the heat released from the hydrolysis and fermentation reactions (both exothermic) were captured simultaneously at temperatures that only allowed for refrigeration as a cold utility. With the SHF technology, heat generated by the hydrolysis reactions was captured at much higher temperatures, where cooling water is used instead of refrigeration. For either case though, external coolers were inserted prior to the SSF/fermentation tanks so that the duties on the refrigeration could be relaxed.

The SSF scenario also demanded more heat for purification, evaporation and boiler water conditioning. This occurred because the feed to the SSF reactor had a solid concentration 12.6%, compared to the feed to the hydrolysis reactor in the SHF setup that had a solids concentration of 13.34, and thus, the excess water impacted on the energy demands of distillation reboilers and evaporators. The marginal increases in the electrical demands reflected the higher flow-rates that were present in the SSF scenario.

The electrical energy for cooling utilities refers to the demands of the cooling tower and the pumping of water from a natural source for cooling. There were no significant differences between the SSF and SHF because the scaling effect of the COP of the cooling tower reduced any significant differences. If it were not for the pinch point retrofits to relax the demand on the cold utility system, then the electrical requirement for the cooling utility would have been 64% greater.

2.6.3.2 **GFT Energy Demands**

Figure 15 shows the specific energy requirements, in terms of MJ per kilogram of bagasse feed, for each of the major process divisions. The highest demand for electrical energy was from oxygen production. Other significant demands for electricity came from raw material handling, compression of syngas for the FT reaction and cold utilities. Majority of the energy associated with the cooling utilities was to satisfy the heat duty of the surface condenser that terminates the expansion circuit. It is in this context where the effect of the pinch point retrofit could be appreciated the most. The total electrical energy demand is 1.41MJ/kg-feed and the pinch point retrofit enabled an additional electricity generation of 0.89MJ/kg-feed. Thus, 62% of the energy needs of the plant were offset, and the equivalent amount was exported as saleable electricity. (Details in APPENDIX 2.2, Table A 4)

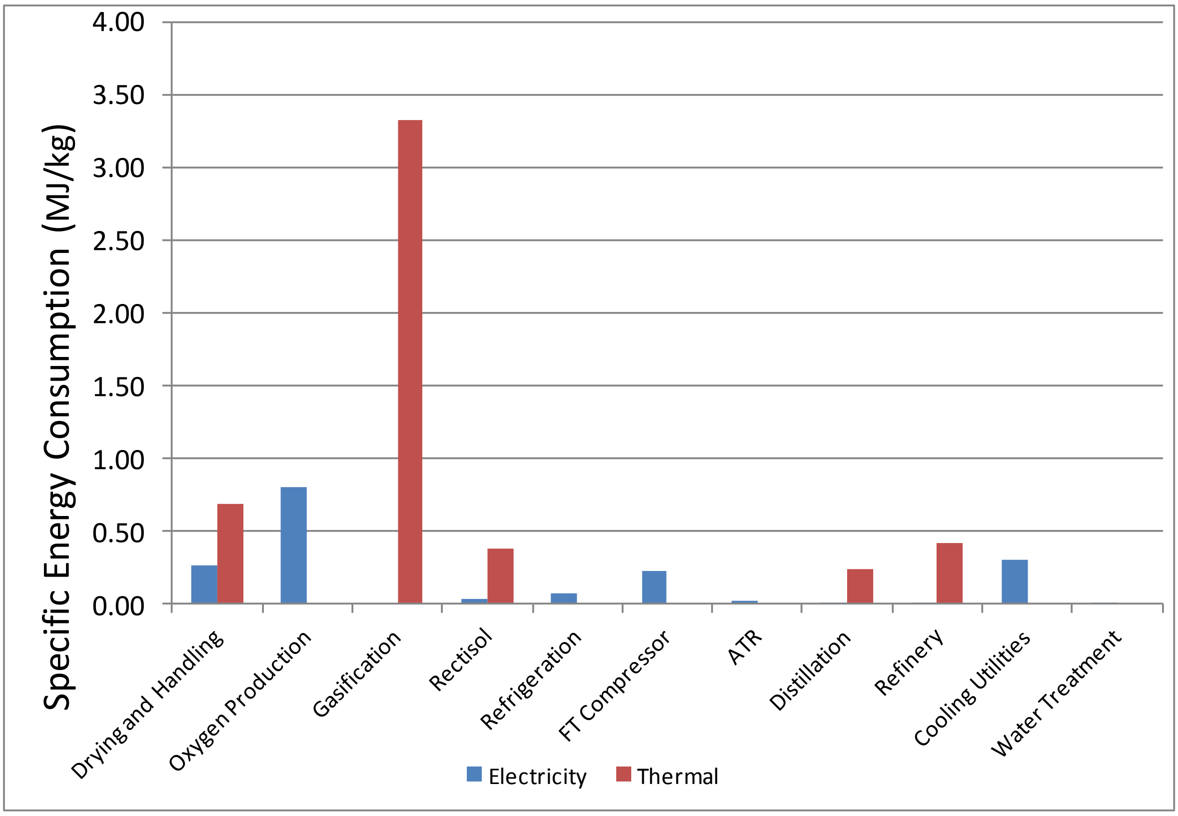


Figure 15: GFT process energy demands

The gasification required 64% of the total heat energy needs of the plant, more than triple the amount of any other heat demand. Other areas requiring significant amounts of heat energy was the Rectisol unit (for methanol regeneration) and distillation with the subsequent refining.

While the drying of raw material does appear to have a high energy demand, it is sufficed by the low quality, high amounts of waste heat generated by the plant (Tijmensen *et al.* 2002). Thus, the heat demand for drying effectively relaxed the cooling utility requirements for the plant.

2.6.4 Process Energy Efficiencies

2.6.4.1 Comparison of Energy Efficiencies of Process Scenarios

Table 6 compares the energy efficiencies of all three processes. The GFT has the highest liquid efficiency, as was expected, since it utilises lignin as feedstock that can also be converted to fuel.

Table 6: Energy Efficiency Comparisons

	SSF	SHF	GFT	GFT **
Liquid Fuel Efficiency	43.6%	38.8%	52.4%	
Total Efficiency	42.8%	39.3%	50.6%	46.4%

GFT** represents GFT efficiencies without additional electricity production.

When comparing the SSF and SHF liquid fuel efficiencies, that of the SSF is only marginally higher. Considering that the design conversion from carbohydrates to ethanol was 68% for SSF compared to 45% for SHF, it was expected that the difference in liquid fuel efficiencies would be in a similar proportions. The reason for the discrepancy is that the recycling system developed by Aden *et al.* (2002), which was subsequently employed Leibbrandt (2010), favoured the SHF configuration only. Such a recycling scheme only allowed a second pass for the unfermented sugars, not for un-hydrolysed cellulose. The losses in the SSF scenario occurred as cellulose being sent for combustion, because the conversion for the singular pass for hydrolysis in the SSF configuration is only 74%, compared to the 83% conversion for hydrolysis in the SHF configuration. If modifications are made to allow this, then the energy efficiency of the SSF process could reach 46% (by calculation).

2.6.4.2 Comparison of Energy Efficiency Results with Literature

Table 7 shows the comparison of the energy efficiencies in this study with values that have been reported in literature. The energy efficiency was either reported in terms of the overall process efficiency or the liquid fuel efficiency, depending on the values that the relevant literature provides. Thus, the overall process efficiency were reported for the SHF and GFT scenarios, and the liquid fuel efficiency for the SSF scenario.

Table 7: Comparison of Energy Efficiencies with Literature

Study Reference	Efficiency (%)
THERMOCHEMICAL MODELS	
Hamelink et al. (2003)	40.00%
Kreutz et al. (2008)	50.30%
Anex et al. (2010)	52.50%
GFT Process	50.56%
GFT** Process	46.42%
Leibbrandt (2010)	50.90%
SHF MODELS	
Anex et al. (2010)	44.10%
Humbird et al. (2011)	47.40%
Leibbrandt (2010)	41.90%
SHF Process	39.29%
SSF MODELS	
Hamelink et al. (2005)	46.00%
SSF Process	43.59%

Humbird *et al.* (2011) obtained the highest efficiency for the SHF scenario, which was primarily due to the optimistic approach that was undertaken in that study with regards to the design specifications of the units. By way of examples, the overall xylan conversion to ethanol was 76% in that study, compared to the conservative conversion of 31% in this study. In addition, the solids loadings to the hydrolysis sections in that study was 20%, while this study assumed 13% based on pilot plant data. This would imply that energy demands of the downstream processes obtained in Humbird *et al.* (2011) would be significantly less, hence further improving the energy efficiency. Similar optimism was applied by Anex *et al.* (2010) regarding the xylan conversions, although the solids concentrations for hydrolysis was more conservative (the details of the biological process in Anex *et al.* (2010) was found in Kazi *et al.* (2010)). With regards to the SSF process scenarios, the efficiency quoted by Hamelinck *et al.* (2005) also exceeded this study due to optimistic approaches. The conservative approach to design specifications in this study was because process conversion specifications were only considered if they were experimentally validated in published literature.

Comparing the energy efficiency of the SHF with those of Leibbrandt (2010) would effectively examine the effect of pinch point retrofits on energy efficiency. It is found though, that the overall efficiency in Leibbrandt (2010) is 2% more and this because the auxiliary energy demands associated with refrigeration and other cold utilities were not taken into account in that study. Thus, the exclusion of utilities demands potentially overestimated the energy efficiency by up to 5% (assuming that the pinch point retrofit increases the efficiency by at least 2.6%). According to Figure 13, these utilities had the most severe demands for electricity.

Considering the superior energy efficiencies of Kreutz *et al.* (2008) and Leibbrandt (2010), to the value reported for the GFT scenario in this study without an additional production of electricity, then a pinch point retrofit does not yield any benefit. Thus, the excess high temperature heat that the pinch point technology avails must be converted into exportable electricity for pinch point retrofits to be of any value in this process.

Anex *et al.* (2010) reported the highest process efficiency for GFT. In that literature however, the description of the process was not detailed enough to determine the cause of the advantage over this study and that of Kreutz *et al.* (2008). However, it is noteworthy because it represents higher potentials for GFT processes. As for the deficit in the energy efficiency of the process of Hamelinck *et al.* (2003), a 50% recycle ratio was assumed there for the light gasses that were sent for Autothermal Reforming (ATR), in comparison to the $\geq$60% specified in this study, Leibbrandt (2010) and Kreutz *et al.* (2008). Reducing the proportion to the ATR would cause an increase in the boiler output and subsequent generation of electricity. This would consequently increase the loss of energy from the entire system since energy losses are generally associated with power generation cycles due to the low efficiencies associated with them.

Thus, the recycle ratio of the light gasses can be considered to be a parameter for optimisation. If the proportion of recycled light gases is increased for better liquid fuel production, then the amount of electricity exported will be minimised. Accordingly, the energy losses from the system would diminish, leading to an improvement the energy efficiency.

The current study only focussed on biological and thermochemical processes and two exclusive scenarios, even though potentials for co-location exist. It has been shown in literature that future technological advances focussing on the merging of the two setups will maximise the energy efficiency (Laser *et al.* (2009)). In such process, the lignin residue from the beer still undergoes gasification and Fisher-Tropsch synthesis to produce FT fuels, while the sugar-rich wastewaters is used to generate biogas to supplement energy needs. Laser *et al.* (2009) has predicted energy efficiencies of up to 79.6% with such technologies. Furthermore, pinch point analysis was not considered in the studies of Laser *et al.* (2003). Given that the GFT process was shown to have high heat excesses (Figure 11), and that the biological scenarios had severe heat demands (Figure 7), the integration of heat between the two co-processes would greatly enhance the energy efficiency.

2.7 CONCLUSIONS

The pinch point analysis of the SHF and SSF processing scenarios has shown any capital expenditure for energy saving becomes necessary because a trade-off between the capital consequences of additional heat transfer area and the economic savings from energy savings could not be reached. In processes that are completely self sufficient, all heat energy that is saved can be converted to additional saleable electricity.

The SSF scenario had superior liquid and overall energy efficiencies, even though the SSF configuration demands much more energy because of excess water and refrigeration. Thus, the greater efficiency is solely due to higher conversions that are possible with an SSF type reaction vessel because the basic flow sheet that was originally developed by Aden *et al.* (2002), only maximising the performance of the SHF process. Measures should be taken to maximise the performance of the SSF process flow sheet as well. One measure to be taken could be to extend the residence time of the SSF reacting vessels to allow a further hydrolysis to occur. The hydrolysis conversion rate in the SHF process should also be achievable in the SSF process, since the pre-treated bagasse had identical digestibility, since the pre-treatment sections were identical for both processes. Such a modification could allow the energy efficiency of the SSF process to reach 46%. Furthermore, the proposed locality was also shown to have favoured the SHF reacting conditions because the sea water would not serve as a adequate cooling medium for the SSF reactions, and consequently refrigeration was needed, which imposed a substantial demand on electrical energy.

Pinch point analysis was proven to have greater beneficial effects on the energy efficiency for the GFT process. Not only has it made high temperature heat available to generate super heat for electricity generation, but provided copious amounts of low temperature steam to achieve the demands of other processes, such drying of raw material. Without the generation of excess electricity, the pinch point analysis would not benefit the GFT process in terms of energy efficiency.

With regard to auxiliary equipment, the electrical energy consumption of the cooling utility units was shown to be significant. Energy balances that were calculated without a reliable estimation of energy needs pertaining to the auxiliary processing steps such as cooling towers and cooling mains pumps could significantly overestimate the energy efficiency.

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3 APPLICABILITY OF FINANCIAL RISK ASSESSMENTS TO BIOFUEL INDUSTRIES

3.1 LITERATURE REVIEW

3.1.1 General Concepts

Irrespective of the technical or environmental feasibility of a proposed undertaking, the decision of whether the undertaking is worthy of a capital investment will depend on whether it is economically viable as well. The economical viability is usually determined by an economic model of some sort. For a processing feasibility, the model would typically be a set of financial statements (Perry and Green (1997)) based on the physical operating data obtained from a mass and energy balance, and capital cost expenditure from sizing the equipment that is involved in the process.

The operating expenses are accounted for in the income statement (Perry and Green (1997)) during the determination of the net profit, and the process data on which the income statement is based is derived from the mass and energy balance. The balance sheet determines the equity by accounting for land values, plant value and liabilities (Perry and Green (1997)). Data for plant value is based on the equipment sizing and material specifications, with some added factors to account for installation costs, and the balance of plant (BOP). The status of the cash reserves is then determined in the cash flow statement by accounting for tax payments and dividends.

The investment decision is based on certain key economic indicators (KEI) which are derived from the financial statements. The KEI of typical use are the net present value (NPV) and Internal rate of return (IRR) (Peters and Timmerhaus (1997)). In an economic model, these values are forecasted over the proposed business life span. The NPV is the sum of the forecasted cash flows over the business lifespan that is discounted at a rate relevant to the industry, less the value of the initial capital investment. Thus, the rule governing an investment decision, with regards to the interpretation of the NPV is that (Balcombe and Smith (1999)):

- If NPV>0, the investment gains returns
- If NPV=0, the investment remains static.
- If NPV<0, the investment decays

The IRR is defined as the maximum possible return rate on the investment and mathematically, it is synonymous to a discount rate at which the NPV remains zero (Peters and Timmerhaus (1997)).

It is common practice though that economic models are largely based on parameters that are averaged , and thus do not account for risks and fluctuations that are realistically associated with the prices of input expenditures and revenues, and with the uncertainties associated with the impacts of mechanical failures (Richardson and Mapp (1976)).

This is true for many techno-economic feasibility studies have been conducted for biofuel prospects. While all these studies have been carried out with sound process models, the economic analysis are based on static prices for commodities, energy and labour, and on economic factors such as interest rates and such as inflation indices. Examples of such instances include:

- Huang *et al.* (2009) investigated the use of aspen wood, stover, hybrid poplar and switch grass to assess which yields the highest techno-feasibility in biofuel production.
- Aden and Foust (2009) who conducted a deterministic techno-economic feasibility study to assess whether the minimum ethanol selling price of $1.37 (2007 USD) can be attainable, using the latest state of technology. The stipulation for feasibility was that the project should break even after 7 years with a 10% discount rate on the projected cash reserves.

In the case of Huang *et al.* (2009) the delivery costs of feedstock were highly aggregated. The delivery cost included agricultural costs that included fluctuating components such as fertilizer replacement cost and farmers compensation. In addition, an average cost was assigned to transport, which inherently has a strong dependence on the fluctuating Basic Fuel Price (BFP) (see Figure 16). Thus, the BFP and other economic variables that display significant fluctuations in their historical trends are considered to be stochastic variables.

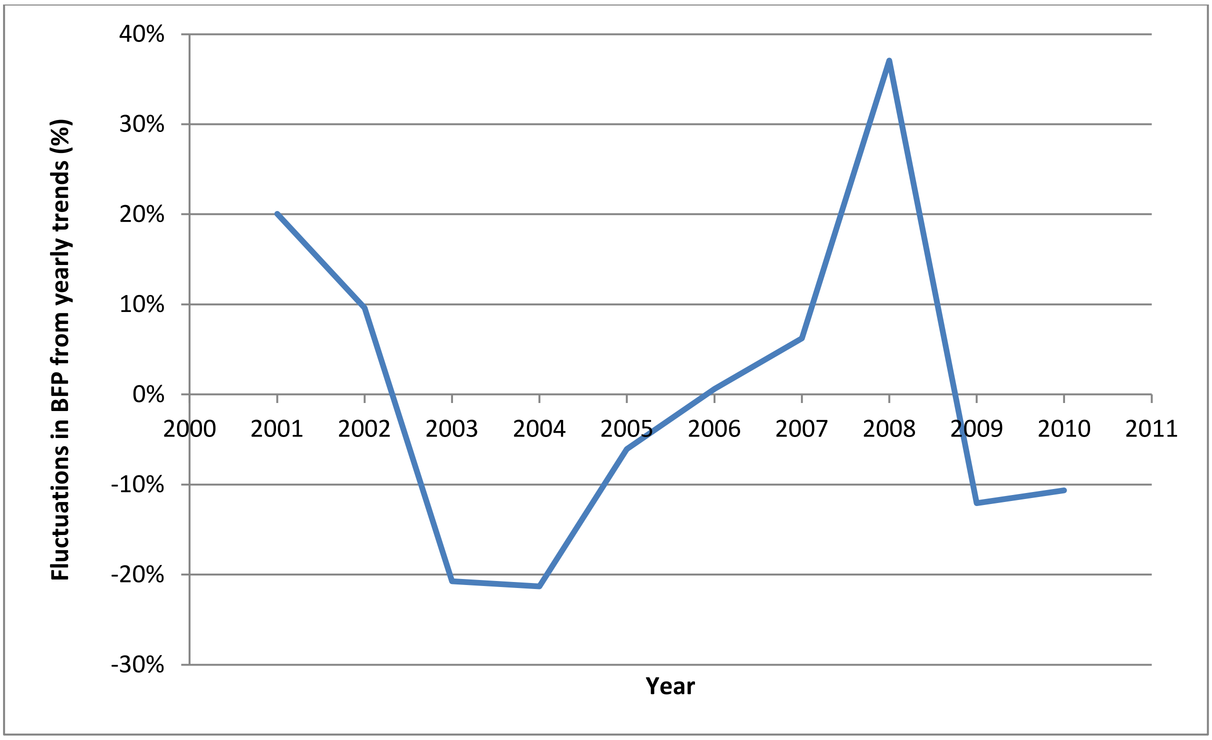

Figure 16: Fluctuations in BFP prices (data from Dept Energy website, RSA)

As with Aden and Foust (2009), major assumptions were made for economic parameters, such as process costs that actually do fluctuate vastly due to global price changes or uncertainty of price determination. Such parameters were often estimated at a mean value. In recognition of these fluctuations however, sensitivity ("what if") studies was conducted on each value to show the effect on the result. This kind of analysis however, only takes into account the extremities in isolation, and do not attach any level of confidence to the result (Oracle 2008). As is shown in Figure 16, the extremities in BFP have very isolated occurrences.

3.1.2 Probabilistic economic models

Instead of using a statistical mean or an assumption for an input, it would be more appropriate that the economic analysis be done in a probabilistic mode. In this way, the question pertaining to the chances of success is fully quantified, rather than predicting a successful outcome that does not quantify risk or uncertainty.

A biofuel plant that derives its feedstock from agricultural sources would essentially be an agribusiness, which means it is liable to encounter the full range of uncertainty and risks that the agricultural industry faces (Richardson et al, (2006)) and that of the petrochemical industry. Such costs would typically be prices of commodities, fuel prices and other operating costs. Even capital costs relating to equipment value are subject to uncertainty because the same equipment imported from various countries do not have the same price (Perry and Green 2008).

In response to the challenge imposed by uncertainty and risk, recent techno-economic evaluations have been carried out using probabilistic economic models to determine the chance of success of bioethanol plants. One such study considered a biofuel plant, where bioethanol was derived from the waste products obtained from wheat harvesting in the Western Cape Region of South Africa (Richardson *et al.* (2007)), and another considered the conversion of corn stover (Richardson and Mapp (2006)) to bioethanol in the Texas region of the USA. Another considered triticale straw as a feedstock (Amigun *et al.* (2011), also in the Western Cape. These studies could potentially enable investment decisions based on quantified risk.

3.1.3 Introduction to Monte Carlo Simulations

The underlying mechanism for the risk assessment models referred to in the section above was Monte Carlo Simulations. Monte Carlo Simulations are processes by which real life problems or phenomena are simulated and a variety of outcomes are expected due to random variations in the input fields (Atanassov and Dimov (2008)). Due to the reliance on repetitive calculations on a domain of randomly generated numbers, the simulations are best performed using computational methods. The focus of this study is the implementation of Monte Carlo Simulation in financial risk assessment for second generation ethanol plants.

According to Moonery (1997), the general methodology for any Monte Carlo Simulation is an easy concept, and can be summarized in the following phases:

1. *Generation of an Input domain*

   A domain is specified which contains the pseudo population that is subject to sampling. The domain is reflected by a probability distribution.
2. *Generation of Random Inputs*

   Generate a random input from the pseudo population which is reflective of the probability distribution.
3. *Compute the outcomes*

   Using the deterministic model or algorithm of the process, determine the outcomes for each random input, and store the outcomes in a vector
4. *Repeat steps 2 and 3*

The sampling strategy and sampling size for each iteration must be consistent.

5. *Aggregation*

   Aggregate the individual outcomes and formulate a distribution that is reflective of the objective.

3.1.4 Application of Monte Carlo Simulation in a Financial Risk Assessment

In a financial undertaking, long term financial objectives require accomplishment under a variety of dynamic factors inherited in human, operating, financial and capital resources (Li and Sun (2008)). While the effects of the current dynamics involved in these constraints can be mitigated through sound management, the unknown future dynamics of the constraints poses risks and uncertainties cannot be determined with concrete certainty.

Considering the rapid developments and changes in the sciences, technologies, socio-economic environments in a projects lifespan, the resulting impacts of risks and uncertainties are subject to rapid changes (Li and Sun (2008)). Thus, it becomes imperative that risk assessment methods are duly integrated in the planning and evaluation phase of a project (Richardson *et al.* (2007)), and be integrated with management practices for continuous attention during the project lifespan (Li and Sun (2008)).

Qualitative risk assessment techniques are traditionally used to account for risk and uncertainty. Commonly used methods are the “what if” scenarios and sensitivity analysis. In these kinds of risk assessments, the optimal, expected and worst case assessments are carried out, but with no indication of the probability attached to the extreme cases occurring (Oracle (2008)).

Risk discount rates are another commonly used method for accounting the risk profile of a prospective investment. A risk factor is used to discount the yearly profits, and this factor is often based on historical data that reflects the riskiness of the prospective investment. The outcome however, does not have a confidence level attached to it (Li and Sun (2008)).

Quantitative risk assessment (aka probabilistic economic models) on the other hand, accounts for uncertainties in expenditure and incomes which are characterized by a probability distribution function. Thus, the outcomes of such an assessment can result in a distribution that can answer questions pertaining to financial success with an established degree of certainty. These sorts of assessments are usually based on the Monte Carlo Simulations.

3.1.5 Generic Layout for a Quantitative Risk Assessment

The general framework for a quantitative risk assessment according to Li and Sun (2008) is, if a consequence (y^j) of a certain risk event (j) is related to the group of randomly fluctuating variables (X_1^j, X_2^j, X_n^j) in a the functional form of

$$y^j = f(X_1^j, X_2^j, X_n^j)$$

then the quantitative assessment of the risks associated with Y is calculated as follows:

1. Generate the domains for each stochastic variable (X_1, X_2, X_n), and generate probability distribution functions $g_1(X_1)$, $g_2(X_2)$… $g_n(X_n)$.
2. Use a computer random number generator to sample each variable in accordance with the probability distributions for each event (j).
3. Use the random numbers of the sampling event (X_1^j, X_2^j, X_n^j) in a risk consequence function (f) to compute this risk outcome y^j.
4. Repeat this procedure for a large number of events, or iterations.
5. Aggregate the outcomes y^j of all events to perform a statistical inference.

Thus, the framework laid by Li and Sun (2008) for a financial quantitative risk assessment is in accordance with the general methodology for a Monte Carlo Simulation which was described by Moonery (1997). In the context of a quantitative risk assessment of a financial undertaking, the consequence (y^j) is evaluated in terms the Key Economic Indicators, which are namely the NPV and IRR (Richardson *et al.* (2007), Amigun *et al.* (2011)).

3.1.6 The use of Key Economic Indicators in a Financial/Economic Quantitative Risk Assessment

The NPV and IRR of an Economic Quantitative risk assessment are derived from a set of financial statements that was dependent on stochastic inputs, which would include the revenues, expenditure and quite possibly, the capital value of the entire project. It was demonstrated by Belli *et al.* (1998) how vastly the fluctuations in the inflation rates can impact on the profitability. Thus, fluctuations of prices associated with basic needs such as water and electricity would have created further uncertainties in the NPV. Furthermore, risk analysis conducted by a tedious series of "what if" scenarios by Tiffany and Eidman (2003) has also shown that bioethanol plant profitability is heavily influenced by market volatility. It was concluded that if there was a repeat in market behavior in the 10 years prior, then the plant would fail to be profitable 8.3% of the time. Statistically expressed, the probability of the NPV being less than zero was 8.3%. The sensitivity was due to the uncertainties in the prices of ethanol, corn and natural gas, or in a general sense, feed, product and utility prices.

Thus, when the aggregated result of NPV's is reached in a quantitative risk assessment (or probabilistic model), as outlined by Li and Sun (2008), a statistical inference can be made in terms of the chance of the NPV remaining positive at the end of the project lifespan (Li and Sun (2008); Hacura *et al.* (2001); James W. Richardson (2006); Amigun *et al.* (2011)). Similarly, with regards to the IRR, the statistical objective would be to determine the probability that the IRR will be greater than a prime interest rate that was guaranteed by a bank (PR). The excepted criteria of a successful outcome is when the chance that NPV<0 is <20% (Hacura *et al.* (2001)). This was a general criteria that was based on previous experience.

With regards to the IRR, it could be expected that the '20%' criteria that is used to assess the NPV be applied to the IRR, if the standard value is assumed to be that of the prime rate (PR). Thus, a feasible investment is reached when Prob(IRR<PR)<20%. The Industrial Development Corporation (IDC) of South Africa (www.idc.co.za/funding/gro-e.asp) however, would fund biofuel projects, such as a second generation plant, with an interest rate of the prime less 3%. Thus, if an investment analysis shows that the IRR has the required probability of exceeding the prime rate, then it would qualify for investment by the IDC. Amigun *et al.* (2011) and Richardson *et al.* (2007) made mention of a minimum private investors return of 25 % for the Return on Investment (ROI). However, a criterion with regards to the probability of the ROI exceeding 25% was not specified, leaving the assessment of such an outcome open ended. Furthermore, the ROI indicator does not account for time-value of money (Peters and Timmerhaus (1997). In terms of the IRR, the incentive for private investment is also 25% (Justice (2009)).

3.1.7 Probability Sub-model for Generation of Stochastic Inputs

The probability sub-model generates the stochastic price inputs using distribution functions and sampling techniques. Thus, the components of this sub-model are the probability distribution function and the sampling technique (explained below). Whichever distribution function is used, Hacura *et al.* (2001) had stressed the need for validating and verifying the distribution function that describes the data and the risk consequence function. Various statistical methods are available for this purpose. Richardson *et al.* (2007) for example, uses the Students T-test to test the ability of the sub-model to predict data sample that has a mean value that is statistically equivalent; and the Box-M Test to test correlation between stochastic variables.

3.1.8 Distribution Functions

Multivariate Empirical Distribution Functions

The simple definition of an empirical distribution function is that it assigns a probability to an event, based on the number of counts leading up to and including the event. Thus, the mathematical formulation:

$$F_n(X_i) = \frac{1}{N} \cdot \sum_{1}^{N} I(x \leq x_i)$$

Equation 3

Where X_i - Element X at incident i
F_n - Probability of element X_i occurs.
n - Number of instances up to and including i.
N - Total Number of incidences
I - Specific Probability Incidences

The empirical distribution is indeed flexible and is especially useful for historical data which does not have a known formal distribution (Kankova and Houda (2006)). Conversely, empirical distributions are also useful for simplifying a problem where the known formal distribution is too complex. When such substitutions are made, the empirical distribution will almost certainly converge to the original distribution (Kankova and Houda (2006)).

Furthermore, Nakamura and Abreu (1993) demonstrated the versatility of the empirical distribution by showing it to have superior functionality than Poisson transforms for such processes that are known to be Poisson. Thus, in the case of a techno-feasibility study that employs a quantified risk assessment model, where stochastic prices and indicators are to be derived from historical data, the empirical distribution would be the most appropriate.

Multivariate Empirical Distributions are used to ensure that the correlation between simulated random variables remains consistent with their historical correlation. This is done by ensuring that the coefficient of correlation remains constant Richardson *et al.* (2000). The method described by Richardson *et al.* (2000) uses an intra-temporal matrix to correlate the variables and a set of inter-temporal matrices for the historical correlation that each variable has. These matrices are used to generate the functions that are used to simulate a multivariate distribution. The same methods were employed by Upadhyay and Young (2005) in determining the risk in an agricultural investment study, when accounting for the risk involved in fluctuating prices and yields. The method for correlating agricultural stock prices by King *et al.* (1988) however, were based on correlation parameters that were derived from percentile values. Comparing that mechanism to the “Richardson method”, the latter can be considered to be more robust and rigorous. This is because the actual fluctuations in market behavior have been correlated in the “Richardson method”, instead of data that had already undergone previously statistical manipulation.

GKRS Functions

The GKRS function is an empirical substitute for the triangular distribution (Richardson *et al.* (2008). It is convenient to use when very little is known about the population other than the minimum, maximum and middle point. In bio-ethanol financial risk models, it has been used to simulate downtime periods (Richardson *et al.* (2007).

3.1.9 Sampling Techniques

The commonly used sampling techniques for Monte Carlo Simulation are (McKay *et al.* (1979)):

Random Sampling

An entire population is classified into a distribution and sampled as a function of a random number and the inverse of the characteristic distribution function. This is by definition, is the most obvious technique for Monte Carlo sampling.

Stratified Sampling

The population is divided up into sub populations that are mutually exclusive, i.e. they contain unique elements of the sampling population. Each sub-population is sampled using the standard random sampling approach, thus ensuring that the entire population is sampled.

Latin Hypercube Sampling

This sampling technique uses a similar approach that the stratified sampling technique uses, except that the distribution itself is discretized. In the context of risk assessment, the number of intervals is equivalent to the number of iterations that are to take place in the simulation. Each interval has an equal probability of being sampled once, and thus the technique ensures that the entire distribution is represented in the input range of samples. Latin hypercube sampling is the most consistent method when compared to random and stratified sampling (McKay *et al.* (1979)), and thus the most reproducible technique.

A concern raised by Olsson and Sandberg (2002), who considered the Latin Hypercube sampling technique for stochastic finite analysis, was that Latin Hypercube sampling can induce a spurious correlation amongst input variables. The problem has been addressed and a simple procedure was developed to overcome the problem.

3.2 METHODOLOGY

3.2.1 General

The data generated by the mass and energy balance will be used to generate a financial risk model. The income statement, balance sheet and cash flow statements will be the base financial statements, from which the data for IRR and NPV calculations will be derived. Thus, the mass and energy balance will provide input to the income statements and the sizing data and costs will be derived from Aspen Icarus for the balance sheet. Refer to APPENDIX 1: Mass Balances and PFD's for data on the mass balance.

Certain elements in the financial statements will be treated as stochastic variable inputs. Data for the stochastic variables over the last 10-20 years will be collected and used to simulate the multivariate empirical distribution. The deviation from trend method will be used for prices (like water and oil), and the deviation from mean method will be used to model economic indices like interest. The steps that form the basics of these methods are (from Richardson *et al.* (2007)):

1. Each price variable and the PPI were modeled with a linear trend-line, and a statistical mean value was used for the bank interest rate and location factors.
2. With each variable, the relative deviation for year was calculated between the data point and modeled value.
3. The relative deviants were then further processed for the multivariate empirical simulations

The stochastic variable distributions will be sampled using the Latin Hypercube Sampling method, to ensure that the entire distribution is represented in the range of inputs. Thus, a total of 500 simulations will be carried out for each process model to generate the required statistical inferences regarding the IRR and NPV. All these simulations will be carried out in MS Excel, with the aid of the Simetar add-on, which is packaged for Monte Carlo Simulations (Richardson *et al.* (2008)). An example of a stochastic input for an iteration of a random variable using the above procedure is presented in Figure 17 below. Note that the data from 2001 to 2010 is the recorded historical data, while the projections from 2011 till 2030 are what the probability sub-Model determines.

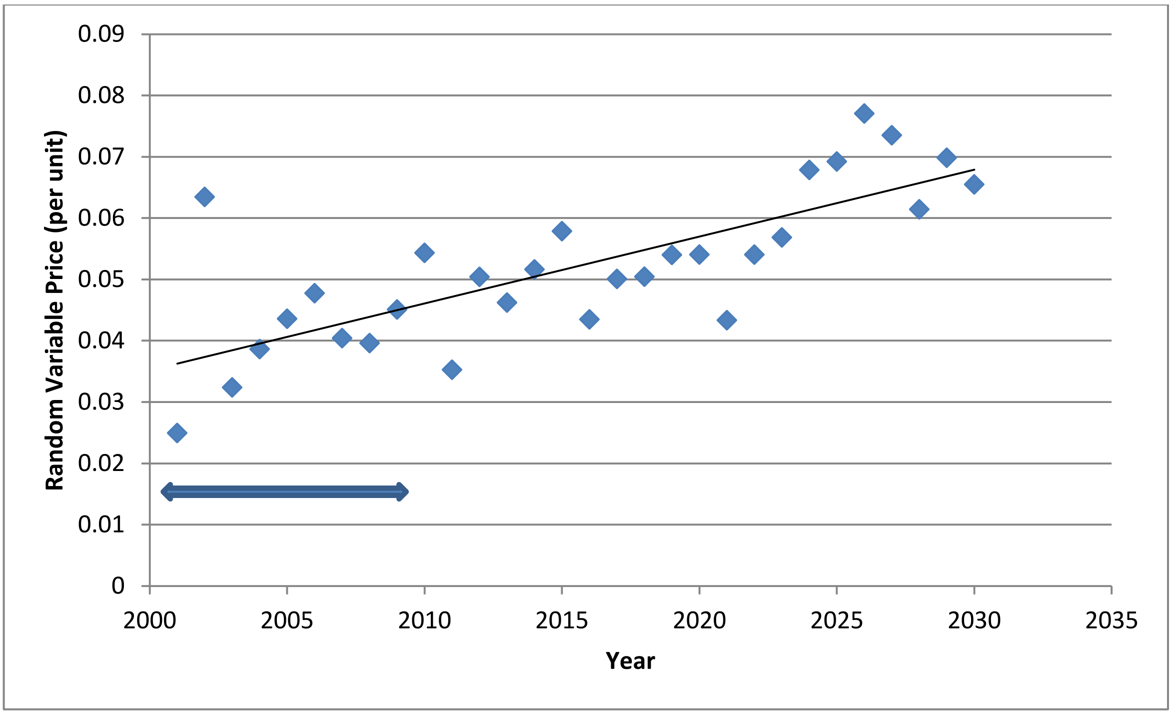

Figure 17: Demonstration of Stochastic Variable Input

3.2.2 Data Collection and Generation for Financial Inputs

3.2.2.1 Capital Costs

Capital estimates were sourced primarily from Aspen Icarus and literature. The application of Aspen Icarus would be very limited though, as Leibbrandt (2010) found that large discrepancies existed between estimates in literature and Aspen Icarus (AspenTech) for identical equipment. Thus, the literature quoted was deemed more trustworthy. In lieu of this, Aspen Icarus was used to size the "minor" equipments, such as heaters, coolers, pumps and compressors, while estimates for more specialised equipment, such as gasification units, fermentation tanks and columns were scaled from quotes in literature. Since the auxiliary equipment such as the waste water treatment plant, cooling towers etc. were not modelled in any detail, capital costs were automatically sourced from literature. Details of the literature derived costs are in APPENDIX 3.1, Table A 8.

In an according manner, the Balance of Plant (BOP) capital costs would be estimated with Aspen Icarus (AspenTech) for the minor equipment types, while literature quoted factors for the BOP was used for the specialised equipment. The BOP costs include piping, civil, instrumentation, painting and insulation. Thus, a correlation was used for estimating the BOP factor, as a function on the scale of the plant (Kreutz *et al.* (2008)).

Thus, the total direct costs (TDC) included the BOP and equipment costs. The total investment costs (TIC) would include the TDC with factored amounts for contingency, engineering, overheads, etc. Factors for these costs were extracted from (Aden *et al.* 2002) for the biological scenarios while it was extracted from Hamelinck *et al.* (2003) for the GFT scenario. Anex *et al.* (2010) relied on (Aden *et al.* 2002) for factors to estimate the indirect costs for the biological scenarios. Regarding the thermochemical scenarios, Anex *et al.* (2010) used factors from Peters and Timmerhaus (1997)**.** The values found in Peters and Timmerhaus (1997) were generally applicable to all operations involving fluid and solid materials, with no specific formulation for the petrochemical industry. Thus, the values found in Hamelinck et al. (2003) were considered because it was more specific to the operations at hand.

3.2.2.2 **Non-Stochastic Operating Costs**

Richardson *et al.* (2007) has shown that there are operating costs that are not stochastic, as they would not contribute significantly on their own to the overall production costs, or attempting to gather historical data for such a large number of variables would be impracticable. These costs could include real estate taxes, maintenance etc. Thus, the values from Richardson *et al.* (2007) were used for the biological models for these parameters. With regards to the GFT process, the overall non-stochastic operating costs were estimated from Kreutz *et al.* (2008).

Flow-rates for chemicals in biological plants were taken directly from the mass balances and while their prices were readily available from the ICIS website (www.icis.com). Even though these expenses were extracted from a static source, they were treated stochastically in a Monte Carlo simulation because the PPI that used to inflate these parameters was stochastically modelled. For a deterministic model, the PPI values are just treated as predicted averages. The cost of enzyme was based on the value of 0.17USD/gal (2007 based currency) that was stipulated in Humbird and Aden (2008). Richardson *et al.* (2007) used a value of 0.045USD/gal (2007 base currency). This value is too low for a conservative estimate, and could yield results that are too optimistic. The list of non-stochastic variables is listed below. The prices and the literature from which the data were sourced from is in APPENDIX 3.2, Table A 8.

- Maintenance & Repair
- Labour

- Management and Qual. Control
- Real Estate Taxes
- Licenses, Fees and Insurance
- Miscellaneous Expenses
- Enzyme
- Fuel Levy
- Non-specific operating expenses for GFT plant

For comparative purposes, catalyst consumption costs for the refinery section and ATR unit was estimated using the pricing data from Silvy (2004). These prices were used in conjunction with the average catalyst consumption rates in European refineries (BREF (2003)) and the spreadsheet model for the refinery calculated from data found in Bechtel_(c) (1997). Furthermore, an estimate for the FT catalyst replacement costs were taken from Kreutz *et al.* (2008) to arrive at a total catalyst replacement cost for the GFT plant. (Note: these costs were used for comparative purposes to put the relative costs of the GFT and biological processes in perspective. These were NOT used directly in the economic models).

3.2.2.3 **Stochastic Inputs**

Most of the stochastic inputs used by Richardson *et al*, (2007) were also deemed necessary as stochastic inputs in the current study. These are:

- Electricity
- Basic Fuel Price (Gasoline BFP and Diesel BFP)
- Feedstock (Bagasse)
- Inflation (PPI)
- Interest Rate

The bioethanol price was determined on an energy equivalent basis with the Basic Fuel Price and the entire fuel levy because bioethanol producers were entitled to receive a full reimbursement of the fuel levy as from 2008 (Biofuels_Strategy (2007)). The feedstock price was based on the energy value of coal, since bagasse could have been used as an alternative feedstock to coal at the boilers (Leibbrandt (2010)).

In addition to these parameters, the following parameters were deemed necessary for stochastic inputs due to variation in their historical databases

- Water (Pricing given by City of UmHlatuze Municipality)
- Diesel Retail

The 10 year price history of each parameter, along with a description of the source, is found in APPENDIX 3.3, Table A 9. The prices will be generated from this database. The difference between the Monte Carlo Assessments and the deterministic assessments are basically that the Monte Carlo forecast uses a predicted yearly price from a linear trend-line AND a stochastic residual, where as the deterministic forecast only uses a predicted yearly price.

Capital costs would have an assumed basis on vendor quotes found in the United States of America (USA). This is true for both Aspen Icarus and literature quotes in the current study. These values were then scaled with the relative ratio of the location factors of the USA and South Africa when the context of the project is localised (USA-RSA cost).

In reality however, a project taking place in South Africa would source capital estimates from a worldwide perspective, and not just the USA. In order to account for this reality, the capital costs would be scaled from a random origin. Thus, a location factor database, using values found in Perry and Green (2008) were used to as an additional stochastic input. This database was generated randomly, with the “random” function found in MS Excel.

3.3 RESULTS

3.3.1 Operating Costs and Revenues

The specific operating cost and revenue, in terms of GJ of feed processed, is shown in Table 8. All these values were obtained for a 600MW plant. With reference to the economy of scale analysis carried out by Leibbrandt (2010), the specific revenues will decrease with smaller scales while the specific operating costs will increase.

This analysis does not form part of the Monte Carlo analysis. With a prior understanding of these economic characteristics of the processes that were modelled, the analysis of the outcomes from the Monte Carlo simulations could be put into context. Moreover, these parameters are somewhat reflective of the mass and energy balances and for that reason could be related as such.

Table 8: Specific Operating Costs and Revenue (USD/kg-feed)

	SHF	SSF	GFT
Revenue	12.83	11.45	13.09
Operating Expense (OPEX)	8.30	7.10	7.98
Cash Income	4.53	4.35	5.11

The specific revenues shown above indicate that the GFT process has the highest potential to produce sales. The ratio of the specific revenues of the GFT to the SSF process is 1.28, which is somewhat reflective of the ratio of the overall energy efficiencies of these processes, which is 1.25. Such an observation further emphasises the necessity to maximise the energy efficiency of a process through pinch point analysis. Retrofitting the GFT process with pinch point had shown that addition electricity was produced at a rate that had offset 60% of the total electrical requirements of the process, thus availing the equivalent amount for sales.

The GFT had specific process costs that were marginally higher than the SSF model. This was to be expected, since the operating expenditure of the GFT included catalysts replacements costs, which amounted USD3.30/kg-feed. This cost by itself was greater than the combined costs of chemicals, labour and maintenance costs of the SSF process.

The SSF scenario had a higher operating expense than the SHF scenario because expenditure on process chemicals were much more. This was due to the excessive nutrient requirements in the Cell Propagation Scheme (AREA 200) of the SSF scenario (SEE APPENDIX 4.1, Table A 11). These additional costs have been shown to increase the overall earnings, the SSF has a better specific cash income.

3.3.2 Data Validation for MVE Simulation in Monte Carlo Analysis

A financial risk assessment model can only be considered valid if the probability sub-model can predict a data set that does not vary significantly to the historical data that the distribution functions were sourced from. As per the method of Richardson *et al.* (2007), the predictive capability of the probability sub-model was first validated. These results are shown in Table 9 below:

Table 9: Probability Sub-model Simulation Validation

		Test Value	**Critical Value**	**P-Value**	**Result**
Test 1	2 Sample Hotelling T^2 Test	5.08	30.15	0.919	*Fail to Reject the Ho that the Mean Vectors are Equal*
Test 2	Box's M Test	40.19	61.66	0.676	*Fail to Reject the Ho that the Covariance Matrices are Equivalent**
Test 3	Complete Homogeneity Test	72.50	72.15	0.047	*Reject the Ho that the Mean Vectors and Covariance Matrices are Equivalent, Respectively**

Thus, none of the null hypothesis regarding the statistical equivalence of the predicted data with the database was rejected. All tests were carried at the 95% confidence. Test 1 implied that the simulated residuals will have an equivalent statistical mean to the historical database and Test 2 implied that the residuals for the various variables were simulated with the statistically equivalent correlation.

3.3.3 Monte Carlo Simulation Results: Capital Cost

Monte Carlo simulation was used to generate capital investment costs as a function of the location of import. Table 10 shows a statistical summary of the results.

Table 10: MC Simulation Results for Total Capital Investment

	GFT	SSF	SHF	Average
Mean (million USD)	813.68	431.81	430.93	
StDev	19.5%	19.4%	19.1%	19.3%
Maximum Saving	38.3%	38.4%	38.4%	38.4%
Maximum Expenditure	32.5%	29.2%	32.6%	31.4%
US-RSA relation to mean	-7.0%	-7.1%	-7.1%	-7.1%

The standard price, which entailed equipment imports solely from the USA, has actually shown to be cheaper than the average scenario, which is indicated by the mean value. Thus, it is 6.95% cheaper to import solely from the USA than from random locations. However, the standard deviation (based on an average) shows that a further 12.84% saving on capital investments could be attained on the USA-RSA price, if the countries from which capital goods are exported are chosen carefully. Conversely, 26.5% can be spent in excess of the standard scenario (i.e. USA-RSA price), if no consideration is given to the country of origin.

Table 11 compares the standard price of the SHF and GFT scenarios with the values obtained in Leibbrandt (2010). All these were based on a 600MW capacity. It was shown by Leibbrandt (2010) that the specific investments, in these scenarios were strongly dependent on scale. As an example, the specific capital investment, per unit of feed processed, for a 145MW GFT scenario was 57% greater than the specific investment of a 600MW scale.

Table 11: Comparison with Base Study

	GFT	SHF
US-RSA (Standard) Price (Million USD)	756.8	400.5
Leibbrandt (2010) Price (Million USD)	732.9	432.9

The results compared very well for the GFT model, with a discrepancy that was less than 5%, even though the approach to calculating BOP and the indirect costs differed - Leibbrandt (2010) used a modelling approach with Aspen Icarus where as this study relied primarily on factors found in literature. Therefore, it can be deduced that the capital breakdown of the equivalent processes will be similar in both studies for the GFT.

The SHF comparison was only reasonably well, the discrepancy being 14%. The specific capital investments from the estimate from Leibbrandt (2010) amounted to USD2430/kW, while that arising from the estimate in this study was USD2031/kW. The estimate of this study also compared reasonably well with the amount derived from the capital estimate of Hamelinck *et al.* (2005) (as determined in Leibbrandt (2010)), which amounted to USD2280/kW.

3.3.4 Monte Carlo Simulation NPV and IRR

Figure 18 depicts the "S-Curve" (Richardson et al, (2007)) that plots the probability of occurrence with of the NPV values from the Monte Carlo simulation. The "S-Curve" can be used to evaluate the outcomes of the NPV from a Monte Carlo simulation in accordance to the stipulation to Hacura *et al.* (2001), which was that a the maximum probability of a zero-to-negative valued NPV is 20% for a feasible investment.

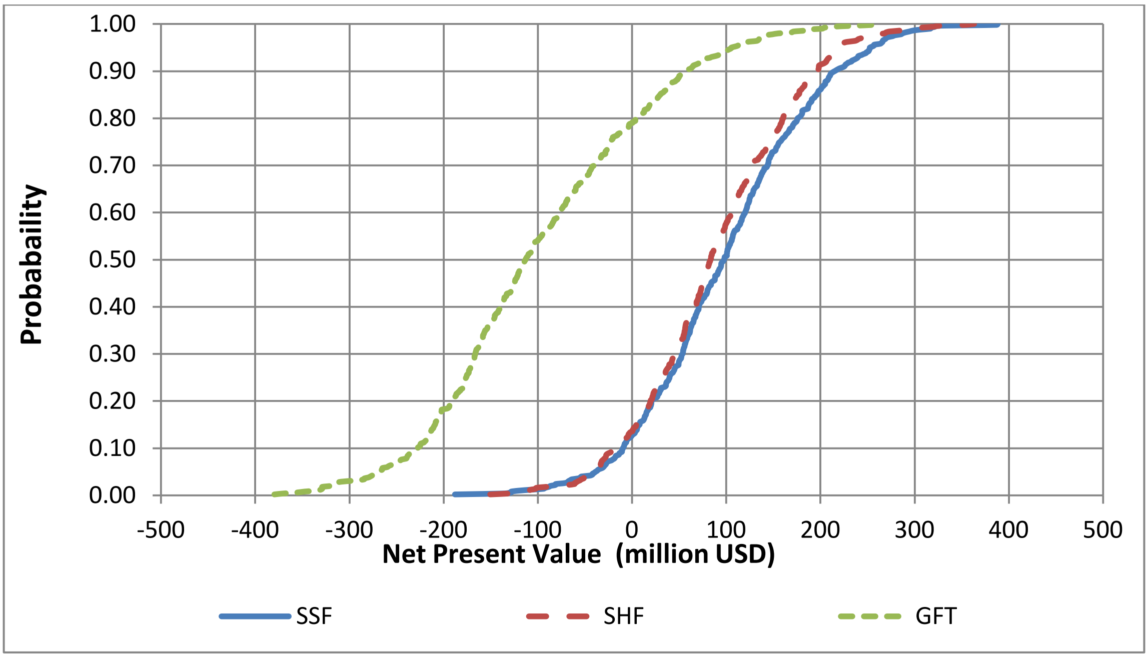

Figure 18: NPV S-Curve

In accordance with that standard, the SSF and SHF scenarios were deemed to be investments of certain feasibility, because the probability of the NPV $\leq$ 0 was 11.5% and 12.1% respectively. Paraphrased, the chance of failure of this investment was 12.01 % and below for either of the biological scenarios. There was only a slight difference in profitability between the two scenarios, because there were no significant differences in the capital investments and net cash potentials between these two processing options.

The GFT scenario showed a far greater tendency towards an economic failure because the probability of a negative valued NPV was 78%. Considering that that the GFT scenario had the best potential to produce a net cash income (Table 8), it shows that economic unfeasibility that was determined for this process was primarily due to the enormity of the initial capital investment. Given that the capital expenditure of the GFT is about double that of the SSF, the net cash potential of the GFT should probably also be double that of the SSF in order to be economically competitive.

With regards to the IRR (Table 12), there is also no significant difference in the outcomes of the SSF and the SHF scenarios. If the benchmark of Hacura *et al.* (2001) regarding the assessment of the NPV is applied here, both these processing scenarios have a similar chance of economic success because the chance of the IRR receding below the bank guaranteed rate (prime rate) of 12.64% was about 10.5% (+/- 0.16%). Once again, the SSF scenario was slightly advantageous over the SHF scenario, because its potential to produce a net cash income was slightly better (Table 8). The GFT scenario still does not show any sign of profitability in the context of the IRR indicator because the probability of receding below the prime is four times that of the maximum threshold. This result is also ascribed of the enormous capital investment. In terms of the 25% IRR standard though, which was described as the incentive for private investors (Justice (2009)) both the biological models would fall short of attaining that standard.

Table 12: IRR outcomes from MC Simulation

	SSF	SHF	GFT
Prime Rate of Bank Guarantee	12.64%		
Probability IRR < Prime Rate	11.31%	12.39%	73.75%
Private Investment Incentive IRR (25%)	25.00%		
Probability IRR < Incentive IRR	81.9%	84.84%	98.94%
IDC Funded Expected IRR (Prime less 3%)	9.64%		
Probability IRR<IDC funded	3.15	3.75	44.82%

3.3.5 Comparison of Deterministic Output to Monte Carlo Output

Table 13 relates the results of the Key Economic Indicators (KEI) that were obtained from a deterministic model to the stochastic model (an actual comparison was only done for the NPV, while the deterministic IRR was just shown, and was compared in a qualitative manner in the discussion that followed). It is shown that across all three of the processing scenarios, the deterministic prediction of the NPV could only be expected with a probability of 33%. This implies that the deterministic prediction of the NPV cannot be assumed to be reflective of the average economic situation, since the probability of its occurrence is significantly less than 50%. One factor contributing to this phenomenon is because the deterministic NPV is based on the USA-RSA capital cost, which has been shown to be 6.9% less than the average capital expectancy of any of the process scenarios that were considered.

Table 13: Comparison of Deterministic and Monte Carlo Outcomes

Variable	SSF	SHF	GFT
Deterministic NPV (USD)	1.33E+08	1.21E+08	-4.6E+07
Prob NPV > Deterministic NVP	34.15%	33.46%	31.80%
Determistic IRR	19.92%	19.33%	10.59%

Regarding the predictive values of the IRR of the deterministic model, the IRR obtained for the biological models was 58% in excess of the prime rate of the bank. Despite that difference, the stochastic analysis has shown that the probability of an unfeasible outcome only receded below the maximum rate of occurrence by 9.34% absolute, and 47% relative (to 20% maximum). This shows the deterministic IRR could be over-optimistic. If the deterministic IRR of the GFT scenario is considered in isolation from the NPV, it exceeded the IDC expected rate of 9.64%, and could consequently be considered as an indication for a feasible investment for the IDC. The stochastic outcome has shown though that an investment in this scenario would certainly result in a failure for the IDC, because the probability of the IRR receding below the IDC rate was 45%, far above the maximum of 20%. Thus, the predictive outcome of an IRR from a deterministic model can potentially be misleading.

3.4 CONCLUSIONS

The SSF scenario showed the best potential for an economically viable investment because the chance of a producing an unfavourable outcome is below the threshold. The outcome of the SSF scenario was only marginally better than that of the SHF scenarios. Investments in these scenarios are very likely to achieve an IRR above the minimum interest rate from a banking investment. The likelihood of this investment obtaining the incentive return of 25% for private investors, however, is still very low. If the SSF process is optimised, then as its energy efficiency approaches that of the GFT, the sales producing potential could reach that of the GFT scenario, and this way it will have a better chance of reaching the incentive for private investors. If it can be confirmed, for example, that the SSF process can operate with a solids concentration that the SHF process operates with, then additional exports of electricity could improve the sales by 6kWhr/MW HHV input, which would amount to 0.36USD/MW_{HHV} of biomass input.

Investments for the GFT scenario would not have a good chance of success because the probability of the NPV obtaining negativity is four times the maximum threshold of 20%. Furthermore, the chance of this scenario producing a return on the investment that is in excess of the minimum IDC rate is not high enough for consideration. This economic un-viability of the GFT process is primarily due to the enormity of the initial capital investment, because the cash producing potential has been found to be far more superior to the SSF scenario. This scenario is likely to be profitable with the n^{th} plant design, since the total capital investment of the n^{th} plant is likely to be half of the current amount, which is reflective on the pioneer plant cost (Swanson *et al.* (2010)). Even so, the biological scenarios would still be more profitable, since the n^{th} plant capital costs for biological scenario is likely to be 73% of the current pioneer prices (Kazi *et al.* (2010)).

Monte Carlo simulations have shown that a straightforward deterministic approach to financial modelling was not a conservative indication of the average expected outcomes. The deterministic NPV, for example, was found only likely to occur with a 33% probability on average, implying that the deterministic NPV was too optimistic. In the case of the GFT scenario, Monte Carlo analysis proved the deterministic outcome false, which showed that the GFT scenario is profitable in terms of an IDC investment. Furthermore, Monte Carlo analysis has shown that capital savings are possible by careful selection of countries where goods are to be exported from.

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4 LIFE CYCLE ASSESSMENT AND THE APPLICATION TO BIOFUEL INDUSTRY

4.1 LITERATURE REVIEW

4.1.1 General Concepts

In promoting sustainability, the environmental impact of a product has to be assessed, whether a service or tangible product. The product that was created in response to a market call has a defined lifespan from the time it is conceptualized and developed, to its raw material extraction, through its manufacturing, transportation, usage, and finally disposal (Ekvall (1999)). Throughout each phase of its lifespan, environmental impacts in the form of emissions, radiation, material and energy consumption, and land usage do occur. Even the conservation efforts such as recycle and reuse, can potentially impact on the environment (Rebitzer *et al.* (2004)).

There are a range of techniques for measuring the impacts that the good/service imposes. These techniques, which are called environmental auditing techniques, are based on input-output analysis, qualitative analysis and substance flow analysis (Rebitzer *et al.* (2004)). They can be applied to existing products to assess the current impacts, or to the design of a new product to evaluate the expected impacts, so that iterative design procedures can take place to minimize impacts (Burgess and Brennan (2001)).

4.1.2 Advantages of Life Cycle Assessments

Life Cycle Assessments (LCA’s) are the most important environmental auditing tool (Burgess and Brennan 2001) to assess the environmental footprint of a product or service. All the direct and indirect environmental impacts associated with a product in its lifespan are comprehensively taken into account and evaluated both quantitatively and qualitatively. The scope of a classical LCA encompasses all the phases of the product life, ranging from the extraction of its raw material right through to the disposal of the product. The design and development phase of a project theoretically form part of an LCA, but it is traditionally excluded, due to the negligible impacts it has in comparison to the latter phases (Rebitzer *et al.* (2004)). Decisions made in this phase, in essence, define the impacts that can be expected in the latter phases of the product life.

Thus, if the objective of the LCA is to ultimately promote sustainability, it should be conducted during the design phase as a complimenting procedure, to assess the environmental viability of decisions that arise from the design and development (Rebitzer *et al.* (2004)). This dynamic was shown in the study of Contreras *et al.* (2009), where LCA was used to determine what the best processing route should be for a sugar cane processing facility. The LCA methodology was used to generate an environmental score for various processing options, which was weighed up against the economic feasibility of each option. Thus, the selected process would have been the theoretical optimum between economic and environmental constraints.

LCAs can form part of a broader optimization tool, as demonstrated by Buddadee *et al.* (2008) to determine the best use of bagasse in Northern Thailand. The model accounted for the environmental optimum through an LCA model and the economic optimum through an economic model. The options under consideration, for bagasse usage were i) combustion for electricity, ii) conversion of bagasse to ethanol fuel or to iii) split between the ethanol conversion and direct combustion. Thus, the LCA methodology was used to score the environmental impact.

According to the ISO 14040 Standard (Guinee *et al.* (1993)) the methodology of an LCA is, in the following order:

1. Framework, Scope and definition
2. Inventory Analysis
3. Impact assessment
4. Interpretation

It was pointed out by Guinee *et al.* (1993), that a 5th phase "improvement analysis" should be included, if the LCA is forming part of an iterative design procedure, or if the project was still in its research phase.

The evident shortcoming of an LCA is that it does not consider the intrinsic risks imposed by a process or operation, as is done through environmental risk assessing (ERA) techniques such as HAZOP analysis. This is because LCA's are concerned primarily with the tangible inputs-outputs of a product system (Burgess and Brennan (2001)).

In contrast to an Environmental Impact Assessment, that will only consider the local impact of a certain economic activity at a particular temporal reference, LCA envelopes the entire range of activities in a products life, and thus potentially evaluating the environmental impact on a global scale. This is however only done if the scope and framework of the LCA is intended for that purpose. Else, LCA's can just as well concentrate on the local impact of the product phase, but with greater quantitative detail. For instance, an LCA can be defined to evaluate and interpret the impact associated with a product from its raw material extraction to usage. The methodology of the LCA, as will be demonstrated later, ensures that the environmental auditing data is collected and processed in a comprehensive manner for logical interpretations. This contrasts the checklist methods employed by EIA's (Burgess and Brennan (2001)).

While Environmental Impact Indicators do quantify the environmental impacts of an activity, pollutant contribution and specific impacts become obscure when indices are formed, and the methods used in formulation of these indices are often highly debatable. LCA's on the other hand quantifies each emission, the impact of the emission, and the consequence on resulting categories such as global warming, acidification etc. Thus, while LCA's do stand as the most superior method of environmental auditing, it should be done in conjunction with ERA's to identify the intrinsic risks of a project (Burgess and Brennan (2001)).

4.1.3 Framework, Scope and definition

4.1.3.1 Definition of a functional unit

The functional unit would be the quantitative description of the service performance of the investigated product system (Rebitzer *et al.* (2004)) that the results will relate to. Thus, the functional unit will often be used in relation to the effectiveness of the product, rather than just the product itself. For example, if the product is motor fuel, the functional unit is better described as the amount of gasoline or diesel used for one kilometer driven in a standard vehicle (Sheehan *et al.(* 2004)).

4.1.3.2 Identification of Processes and units to be included in Framework

Though the idea of an LCA was originally to follow the impacts of a product from the origin (raw material extraction) to its end (disposal), the analysis can become impossible without system boundaries (Burgess and Brennan (2001)). In addition, the acquisition of data for a complete LCA can be unobtainable because such data records could not exist, or parties involve would be unwilling to divulge information (Jimnez-Gonzfilez *et al.* (2000)). Thus, assumptions become necessary, and the number or assumptions, or the broadness of a particular assumption will require weighting the completeness of the study, against the practical feasibility of the study (Guinee *et al.* (1993)).

Processes considered in the system are those deemed to have a significant contribution to the functional unit of the product, and the emissions to the environment (Guinee *et al.* (1993)).

Whether the processes involved in the production of equipment and capital goods should be included in the study depends on the durability of the goods. If mechanical equipment, for example, needs constant replacement or maintenance, then the framework should include those processes to account for the environmental burdens associated with the production of such equipment (Frischknecht *et al.* (2007). Furthermore, if the production of the capital goods uses significant toxic substances, such as solar panels, then the boundaries should include the production process of the capital goods (Frischknecht *et al.* (2007**)**.

The objective of the LCA will strongly influence the positioning of the system boundaries. For example, if a comparative study is conducted on a product to compare the impacts associated with alternative processing, or manufacturing routes, then the equivalent phases can be omitted in order to focus on the processes in question (Burgess and Brennan 2001). While conducting comparative LCA studies however, biasness should not be created in the placement of system boundaries, which is evident from the study of Niederl and Narodoslawsky (2004) who conducted a comparative LCA of a biodiesel production, either using tallow or spent vegetable oil (SVO) as competing feedstock. The extending of the system boundaries for evaluating the tallow option to include the cattle farming created a biasness that was in favour of the SVO, as the boundary for the SVO were fixed at the facilities that were producing the SVO as waste. A fair comparison would include the agricultural activities for producing fresh oil as well. Indeed, it has been shown in an LCA study of bioethanol production from corn-stover (lignocelluloses) (Sheehan *et al.* 2004) that the agricultural activities for harvesting stover had a significant contribution to the overall impact.

4.1.3.3 **Process Orientated Life Cycle Analysis (Process Environmental Assessment)**

Application of LCA's has mainly focused on products, i.e. Product Orientated LCAs. The usage of Process Orientated LCAs to optimize, or at least contextualize the environmental burdens of a process has been limited (Burgess and Brennan 2001). The procedure for Process Orientated LCAs, otherwise called "gate to gate", or Process Environmental Assessments (PEA's), has been outlined in Jimnez-Gonzfilez *et al.* (2000). PEA's can compare alternative processing technologies (Melamu 2008) in terms of their environmental impacts. Otherwise, the results can be interpreted in terms of the specific process design to locate the deficiencies that were the cause of excessive environmental impacts. Such actions could be iterative to arrive at the best design for environmental sustainability. Thus, it is more likely that the type of LCA that was referred to in the "LCA-Process Design" procedure by Rebitzer *et al.* (2004) is the Process Orientated LCA (aka PEA's).

The allocation procedure for PEA's has not been clarified yet (Burgess et al, 2001). It does not seem appropriate to distribute impacts between products in PEA's since it's the processing technologies itself that is of concern. Thus, a functional unit for comparison is likely to relate to a reference flow describing the feedstock characteristics. In a study to compare the environmental impacts of processing sugars and starches either by ethanol or hydrogen production (Melamu 2008), a reference flow of 1 hectare of agricultural land was employed as the basis of comparison.

4.1.4 Life Cycle Inventory

The Life Cycle Inventory (LCI) analysis of an LCA quantifies all environmental flows associated across the system boundary defined in the initial phase of the LCA (Guinee *et al.* 1993). If the LCA concluded at this stage by using the information of the inventory to minimise the impacts, it would essentially be rendered as a detailed EIA, because the qualification of the relative importance of emissions in terms of the category of environmental burdens is not determined (Rebitzer *et al.* 2004).

4.1.4.1 **Meta Data**

Data documentation which describes the geographic, temporal and technological validity of the data is sometimes deemed to be as crucial as the data itself (Rebitzer *et al.* 2004). Thus, the ISO14048 standard specifies that support documentation should be structured around:

- Process description to explain all environmental flows (in essence a technological description (Burgess and Brennan (2001)).
- Modelling and Validation of data.
- Administrative information, regarding the temporal and geographical importance.

However, Burgess and Brennan (2001) point out that the importance of geographical information is not of importance in a Product oriented LCA, as the primary objective of an LCA is to promote global sustainability and that temporal validation is not necessary, unless there is strong evidence to show that yearly throughputs have vast differences.

4.1.4.2 **Co-products of a process**

When dealing with multi-output processes, the environmental burdens must be distributed proportionally between the products. However, the standards or parameter to define the proportionality, or allocations, is unclear, and various LCA's have considered various parameters on which to base the allocation (Kodera (2007))

Commonly used parameters are mass, volume, energy and economic value. Economic value as an allocation bases is increasingly being promoted for the following reasons:

- Different physical parameters can yield vastly different outcomes (Guinee *et al.* (1993)).
- The economic value of the product or the process economic feasibility is the primary justification for the project existence (Kodera (2007)).

"Cause and effect" is an important consideration, as emissions can occur upstream from the point on the production chain where a co-product is produced. In that case, the one product will not share in environmental burdens associated with dedicated chain of the other co-product (Burgess and Brennan (2001)).

Another method for accounting for co-products is by crediting the primary product with the environmental burdens that will insure from a "dedicated" process that delivers the co-product as the primary product (Sarantakos *et al*. (2008)), i.e. substitution. However, this procedure will require substantial data collection for the inventory, as the system boundaries needs expanding to incorporate all the dedicated process (Burgess and Brennan (2001)). The result of such an LCA becomes blurred and cannot be interpreted easily. An example of such blurring is demonstrated by Niederl and Narodoslawsky (2004). When comparing the results obtained by substitution against that obtained by allocation for the glycerol bi-product, the substitution results were vastly lower, having negative values.

Biofuel plants usually produce some sort of co-product, such as glycerin from biodiesel plants and electricity from bio-ethanol plants. Majority of the cited examples in literature used the substitution method to account for co-products. It was shown to yield results that are often blurred and not easily interpreted when comparative studies are done. It is not surprising though that many authors follow the substitution method, since it is the recommendation of the EPA-USA and the ISO14040. However, both UNFCCC (United Nations Framework Convention on Climate Change) and IEA

(International Energy Agency) have advised the counter, because of the blurring of results that arises from substitution (Sarantakos *et al.* 2008).

There is a risk that the substituted product is not compatible in terms of its potential to deliver the function that is required. This sort of incompatibility was shown in Contreras *et al.* (2009) where a fertilizer was used to substitute for a filter cake that was co-product of an ethanol production process. Though the equivalent amounts of fertilizer were determined of a nutrient level balance, the assimilation rates of the nutrients in the commercial fertilizer would obviously be much more rapid than that of nutrients locked in an organic state in the filter cake. This deficiency of the filter cake as a fertilizer was not accounted for in the study. Furthermore, biogas, another co-product was substituted with petroleum kerosene, and subsequently, the impacts associated with kerosene production from petroleum refinement were credited from the ethanol process. This substitution does not seem completely compatible as kerosene arises as co-product of gasoline and diesel production. This does not conform to the ISO standard, which prescribes that a process used for the substitution of a co-product, should have its co-product as its primary product.

4.1.4.3 **Inventory Data from Mass Balances**

Challenges in data acquisition are often due to the response time from the parties to whom the data belongs. As pointed out by Rebitzer *et al.* (2004) and Jimnez-Gonzfilez *et al.* (2000), this is often due to concerns in terms of intellectual property or fear of exposure. Else, the emission data was recorded for organisational purposes, rather than for functional purposes, and the resulting quality of the data reflects the lack of detail. One route of overcoming these concerns is the use of industrial average's found in literature or published databases, which will inherently dampen the level of confidence to the project. Otherwise, a more robust method will be to carry out a mass and energy balance using the standard processing engineering techniques (Jimnez-Gonzfilez *et al.* 2000). Other than the obvious advantage of confident data, the data is validated through modelling, and the exercise will be accompanied by a process description. Therefore, it will inherently fulfil the requirement of the ISO14048 standard for data support documentation (Jimnez-Gonzfilez *et al.* (2000)).

Jimnez-Gonzfilez *et al.* (2000) outlined the standard procedure for completing a mass and energy balance. A validated computer simulation of a mass and energy balance will yield data with better accuracy. It should be stressed that the data yielded by a mass and energy balance be reconciled with measured data because assumptions are likely to be made (if measurements are available). As an example of such an assumption, it is specified that fugitive losses for liquids with boiling points between 20 and 60°C should conservatively be estimated at 2% while those with higher boiling points are estimated at 1% (Jimnez-Gonzfilez *et al.* (2000)).

Computational procedures were shown in the study of Sheehan *et al.* (2004), who conducted an LCA of bioethanol from corn stover, to produce an LCI of high quality that met all the specified standards. Inventories of all processes considered in this system were derived from modeling studies. Sophisticated agricultural models were used to determine the emissions from the harvesting and soil emissions and mass and energy balance models was used for the processing of corn stover to Bioethanol.

4.1.5 Life Cycle Impact Assessment

4.1.5.1 **Impact Categories**

The ISO 14042 (Pennington *et al.* (2004)) stipulates that the three broad categories of impacts that should be accounted for when defining the scope of an LCA are:

- Resource use
- Human Health Consequences
- Ecological Consequences

Within the broad categories, are the specific impact categories. These, together with their respective unit of measurements, are listed in Table 14 (data extracted from Pennington *et al.* (2004))

Table 14: Impact Categories of LCA

Specific Impact	Unit of Measure
Climate change (Global Warming)	Global Warming Potentials (unit mass of CO_2, or the equivalent)
Stratospheric ozone depletion (ozone destruction)	Ozone depletion Potential (1 unit of CFC, or equivalent)
Acidification (contribute to acidifying rain)	Hydronium ions formed per unit of an acidic emission. (mass of Sulphur dioxide or equivalent)
Aquatic Eutrophication (nutrient enrichment of water body causing growth of Algae)	mass or equivalent of phosphorous
Terrestrial Eutrophication (nutrient enrichment of soil causing growth of alien plants)	mass units of nitrogen
Human toxicological effects (detrimental impact on human health)	Typically characterised by the benzene equivalent, which is thus only considering carcinogenic substances. Recent studies (Pennington et al, 2002), have aimed at expanding toxicological effects in an LCA to be inclusive of non-carcinogenic toxic substances.
Ecotoxicological effects (destroys fauna/flora)	Mass of dichlorobenzene or equivalent.
Photochemical oxidant formation (free-radical formation)	Measured as non-methane volatile organic carbons
Biotic Resource Depletion (Fauna/flora resources)	Over exploitation can lead to detrimental effects on diversity and even extinction.
Abiotic Resource Depletion (Fossil and Mineral Resources)	extraction rates
Continuous land use impacts	area of use multiplied by the time of duration
Land transformation	area change

The impacts of the processed data from an LCI are typically quantified in one or all of the above categories to determine the nature of the environmental burden (Pennington et al, (2004)). This could be useful to assess the relative impact levels when comparing alternative processes for an identical product. Complications arise when the normalisation is considered and the when geographical settings offer different sensitivities to the impacts.

From a review of LCAs conducted for biofuel production (Buddadee *et al.* (2008), Sheehan *et al.* (2004), Niederl and Narodoslawsky (2004)), the following categories have been significant:

- Global Warming Potential
- Acidification
- Abiotic Depletion
- Eutrophication
- Human Toxicity

4.1.5.2 **Method of Analysis**

The method of analysis refers to a publicized system of calculations that is used for conducting the impact assessments from the inventory data. There are a number of methods that have been published, some suitable for specific types of applications. One such method recommended for use in bio-energy based LCAs, according to Melamu (2008), is the CML 2 Baseline 2000 v2.05. This method incorporates all the impact categories deemed important for a biofuel based LCA and it fulfills all ISO standards (Melamu 2008).

4.1.5.3 **Normalisation**

Normalisation of impacts, though not compulsory, is deemed important in an LCA, as it leads to decision making pertaining to trade-offs between the various categories. However, the major concern around normalisation is that there is no standard scientific methodology for formulating the weighting factors to normalise the categories (Finnveden (1999)). Thus, as with environmental impact indices, the suggestive use of “expert panels” are employed to qualitatively developed the weighting factors. Problems arising from expert panels vary, from those associated with biasness to the types of expertise to be found in the panel. Finnveden (1999) also pointed out the output resulting from panels methods would vary with the manner in which identical problems were postulated.

Even though weighting factors can be based on political and scientific standards or goals, the relevancy only applies for as long as the standard applies (Burgess and Brennan (2001)). Thus, if an environmental standard regarding emission level changes, a new normalising result can yield a vastly different conclusion than the first. A further weakness, as highlighted in Burgess and Brennan (2001), is the weighting factors of critical levels (i.e. point of irreversible damage) are given the same weighting across all categories of impacts. Assignment of weighting factors could potentially ignore the emphasis on impacts that are more crucial to the geographical setting.

An even further weakness, such as irrelevance to certain impact categories, is shown by the normalisation in the study of Niederl and Narodoslawsky (2004) using an index called the 'SPI' (sustainable process index). This index is based on the assumption that the economy builds on the maximum amount of work that can be derived from solar radiation. Such an index would have no clear relation to impact categories such as eutrophication and land toxicology. On the contrary, the study of Sheehan *et al.* (2004) did not use any normalisation techniques, and therefore could clearly show that contributions to impact categories of acidification and photochemical smog from biofuel production and usage were higher than the impacts arising from equivalent amounts of fossil fuel processing and combustion .

4.1.5.4 **Temporal Considerations**

LCA's are static processes which imply that environmental flows can be averaged at a steady flow. However, the steady state flows can have dynamic effects on the environment (Udo-de-Haes *et al.* (1999)). If there isn't an outflow from the ecosystem, into which the emission occurs, a build-up of pollutant will occur that can eventually lead up to pollutant levels that surpass the critical value. In addition, the effect of the impact on the environment can have a dynamic effect, for example, phosphate discharges into a fresh water body will cause an algal bloom long after the phosphate emission has taken place. Such dynamic effects are not modelled in SIMA-PRO.

4.1.5.5 **Geographic Considerations**

Certain environments have a natural absorption capacity to a certain impact, but can be very sensitive to another (Udo-de-Haes *et al.* (1999), Burgess and Brennan (2001)). For example, CO_2 emissions in a rainforest will be buffered by the carbon uptake of the vegetation.

4.1.6 Review of LCAs conducted to promote Sustainable Energy

A comparative study for determining an environmental and economic optimum processing routes for sugar cane was carried out by Luo *et al.* (2009). A comparative study for determining the environmental and economic optimum processing routes for sugar cane was carried out by Luo *et al.* (2009). The two processing routes which are compared here are a first generation ethanol plant and a hybrid first-and-second generation plant. Thus, the first generation plant was considered as the base case where 48% of the sugarcane juice was portioned towards ethanol production, while the bagasse is burnt for electricity for the both the sugar and ethanol plants. The remaining sugarcane juice was used for sucrose production. The hybrid first-and-second generation plant was treated as a future case where the bagasse is used to supplement ethanol production.

In contrast to the other studies mentioned, Luo *et al.* (2009) used economic allocation instead of substitution for co-products, and since no “negative impacts” were present, the study remained completely transparent. Across most of the impact categories, the future case (higher ethanol production) had lower environmental impacts, and it performed better financially. The future option had the highest global warming emissions though, because the combined energy needs for both the ethanol and sugar processes were not fulfilled by the amount of bagasse that was combusted for energy. Therefore, additional fossil energy was required.

In Tanzania and similar developing states, the reliance of fuel oil for domestic use is still quite intense. Thus, fast pyrolysis of wood to produce bio-oils for such purposes would serve as an attractive alternative to reduce the negative impacts (and reliance) associated with fuel oil production and combustion. An LCA study was conducted by Manyele (2007) to determine the prospective environmental sustainability that can be expected from fast pyrolysis of wood. The system boundaries encompassed the entire production life cycle, from the plantation of perennial vegetation for feed stock, to the complete end of the product life where bio-oil is combusted. Inventory data was obtained from mass and energy balances, and certain well established industrial mean values. Impacts were categorized into human health, environmental health and land usage. While the impacts on human and environmental health categories remained low, effects on land sustainability can be very adverse if natural land is used for plantations. Thus, if proper land use and feedstock generation strategies are in place, such as plantations on existing agricultural land that is not in use, and strategies to avoid biomass at landfill sites are in place, then land sustainability can also be promoted. Air emissions such as SO_2 from pyrolysis oil combustion are very low when compared to fuel oil combustion. Nitrous oxide levels can vary though, as nitrogen in biomass varies. Thus, the feasibility and environmental promotion of pyrolysis can vary depending on the feed.

In complementation to this observation, which was that environmental impacts of a pyrolysis process can vary depending on the feed, Roberts *et al.* (2010,) in an independent study compared the LCA of a slow pyrolysis process in response to three alternative feedstock types: stover, switch grass and yard waste. Boundaries for the stover were set at stover harvesting (compare Sheehan *et al.* (2004)) and thus, inventory data for stover harvesting were obtained from LCA's conducted specifically for that. Since switch grass is planted for bioenergetics reasons, the boundary for this scenario included the impacts associated with all the agricultural activities. Regarding the yard waste scenario however, the boundaries was expanded to accredit this processing scenario with the composting processes that alternatively uses the yard waste as a feedstock. This was an invalid substitution by definition, because the yard waste is just a feedstock and not a co-product. With regards to the performances of the types of biochar resulting from the various feed stocks, if the performance can be defined as the potential for the most efficient sequestration of stable carbon, the stover produced biochar is the most superior.

As to whether pyrolysis and gasification is superior to direct incineration, Soft RJ (2007) conducted a comparative to compare two alternative processes for creating sustainable energy from paint waste. One process considered the production of thermal energy through pyrolysis, gasification and combustion to treat contaminated soil. The other process considered direct incineration to provide thermal energy for electricity generation and distilled water production. For either process, the system boundaries were taken at the gate of the waste storage to the production of the final products, which was decontaminated soil in the first scenario and electricity and distilled water in the other. Inventory for all the processes considered in both systems was obtained by mass balancing, and conservative estimates were made during the mass and energy balances so that data is not to optimistic. The impacts were assessed in a wide range of impact categories and indicated that both processes have a high impact potential on air quality, and toxicological effects on water and land systems. This is expected though since paint wastes are extremely toxic themselves and would have a far greater impact if disposed of in conventional methods.

Thus, a comparison of pyrolysis/gasification to incineration proves that the latter has much more environmental burdening than the former (Soft RJ (2007)). This result concurs with the theoretical prediction that pyrolysis/gasification will have better energy efficiency and lower emissions. In direct combustion, there are ash layers which limit heat and mass transfer. These limitations can be overcome with a high gas velocity (Grishin and Yakimov (2009)). Pyrolysis and gasification reduces the substances into simple gases, and thus, these limitations are absent during combustion. Thus, the exhaust gas for scrubbing would have occurred at a much smaller flow rate than the equivalent exhaust gas resulting from direct combustion, which means it can be cleaned easier in a scrubber.

4.2 METHODOLOGY

4.2.1 Life Cycle Methodology

The methodology will be in accordance with the key recommendations found in the literature survey. To summarise these findings with each phase involved in an LCA:

1. Goal and Scope
    - Use attributional LCAs for each process.
    - "Loading to Grave" boundaries for product oriented LCA.
    - "Gate to gate" boundary for the PEA.
    - Functional unit defined as the fuel required for 1km driven by a standard vehicle for product orientated LCA.
    - Basis of comparison is 600MW of bagasse for PEA.
2. Inventory
    - Simulate data wherever possible, then after, resort to data sources.
    - Allocation by economic revenue for product orientated LCA.
    - Neither allocation, nor substitution required for PEA.
3. Impact categories to be considered:

    - Global Warming Potential
    - Acidification
    - Abiotic Depletion
    - Eutrophication
    - Human Toxicity

    Avoid normalising across impact categories

4.2.2 Software Used

i. SIMA-PRO (Pre-Consultants (2010)) – formulated to conduct Life Cycle Analysis for a project. Used by Contreras *et al.* (2009), Melamu (2008) and Buddadee *et al*. (2008)

ii. GREET 2.7 (Wang *et al.* (2007)) – use to calculate emissions from engines and various processes relating to the motor and fuel industry. Used by Melamu (2008).

4.3 RESULTS

4.3.1 Goal and Scope

4.3.1.1 Definition of System Boundary

LCA studies were conducted on the biological process routes and the thermochemical route for comparative studies. Thus, the LCA’s were of the attributional process type, with a system defined from “loading to grave”. Therefore, the production of the raw material, namely the bagasse, is not considered in this exercise because the entire scope of the study is to find the best option for transport fuel derived from the excess bagasse. The impacts from each process within the boundary, along with all environmental flows across system boundaries are considered in the overall impact assessment. A schematic diagram for the general system is depicted in Figure 19.

For the process orientated LCA, the boundaries simply exist around the production facility. Thus, impacts from the production process, along with all environmental flows across its boundaries are considered in the overall impact assessment.

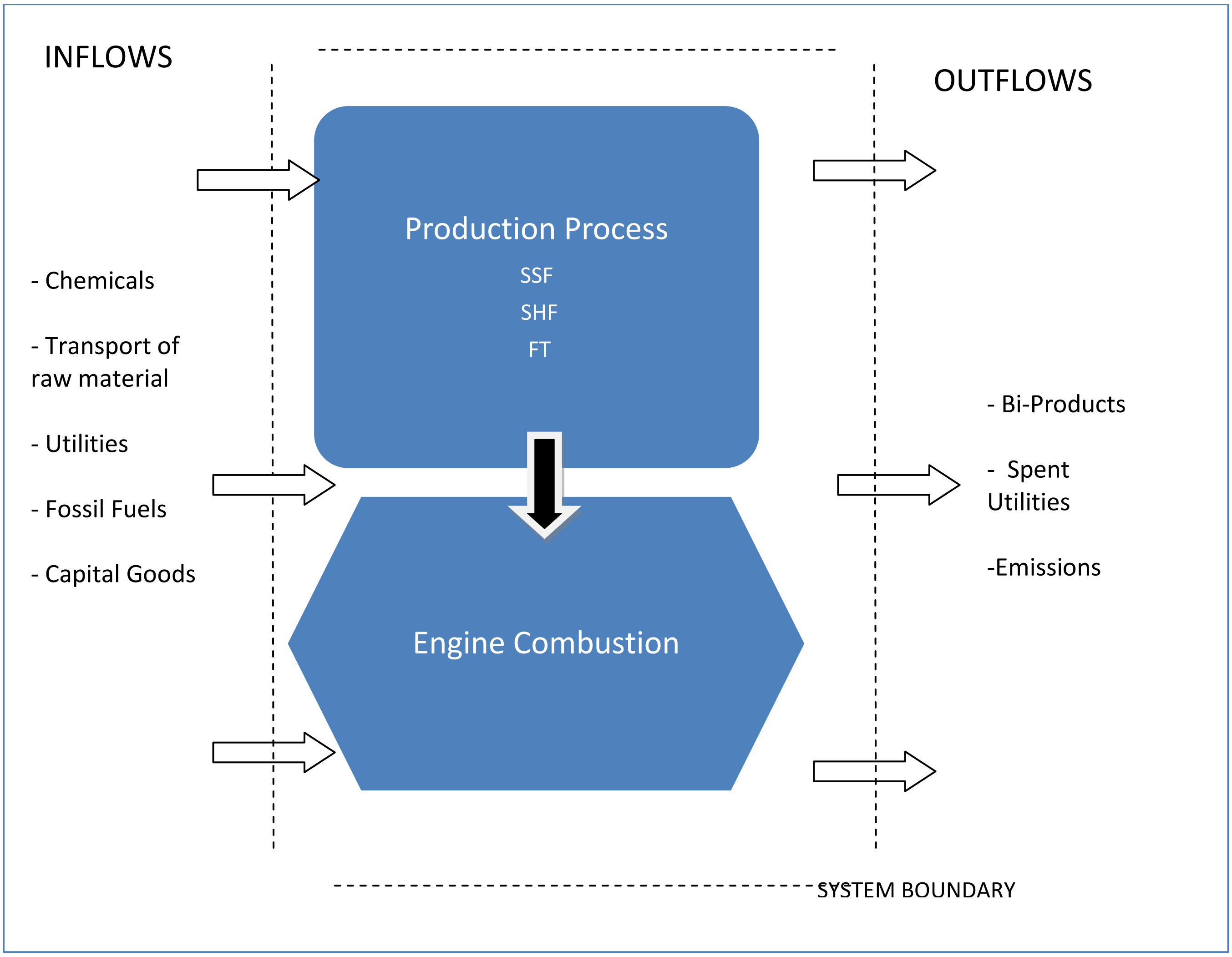

Figure 19: SCHEMATIC DIAGRAM OF SYSTEMS FOR LCA

4.3.2 Functional Unit

The quantity of fuel that an engine required to deliver a standard motor vehicle over a distance of 1 kilometer served as the functional unit for a product orientated LCA. From the GREET 2.7 model (Wang *et al.* (2007)), it was calculated that the equivalent amounts of ethanol fuel and gasoline were 136grams and 88grams respectively. Thus, all processes and flows were automatically scaled in SIMA-PRO accordingly to meet the delivery of these amounts of the respective fuels.

For the process orientated LCAs, which focused on technology comparisons rather than product comparisons, the functional unit, or more appropriately called, the basis for comparison was the thermal input of 600MW of bagasse processed (corresponding to 113.7 t/hr of dry biomass).

4.3.3 Life Cycle Inventory

4.3.3.1 Report of Data Acquisition and Generation

Inventory data for the processes scenarios was generated by means of the mass and energy models in Aspen Plus, with complimentary MS Excel spreadsheets to model processes that were not simulated in Aspen Plus. Inventory data was also sourced from industrial averaged emission measurements in literature and in databases within software such as GREET 2.7 (Wang *et al.* (2007)) and SIMA-PRO (PRé-Consultants 2010). Process descriptions accompanying the mass and energy balance fulfill the meta data requirements of the ISO14048 (Rebitzer *et al.* (2004)).

Inventory data would generally fall within two categories: process inputs and emissions. Process inputs refer to all material and energetic inputs that flow inwards across the system boundary such as raw materials, chemicals and utility flows. Data for these parameters were derived exclusively from the mass and energy balances from Aspen Plus models with complimentary calculations carried out with MS Excel. Material transport, which was considered as an input, was modeled within SIMA-PRO. All input data for the processes that were generated for the LCI is given in APPENDIX 4.1.

Emission data for the LCI were all process outflows besides the primary and co-products (Rebitzer *et al.* 2004). The data was gathered from multiple sources. For the biological scenarios, the data generated by the mass and energy balance usually sufficed. With the GFT scenario, the data generated by mass and energy balance models needed modification to some extent to reflect reality because of the numerous amounts of constituents that are present in refineries. The mass balance of the refinery of the GFT scenario was used in conjunction with the values reported in the BREF (2003) to generate emission data for the inventory. The emissions associated with processing units in the refinery section was reported in the BREF (2003) represented the averaged values for refineries in Europe. Thus, the resulting emission inventory had a quality standard in-between the high standards associated with mass balancing and the lower standard associated industrial averaged data. Emission data from the processes that were generated for the for the LCI is presented in APPENDIX 4.2

Only the emission inventory describing the constituent data of air emissions were considered because water quality constituents were accounted for by built in water-treatment models within SIMA-PRO (Ecoivent (2007)). The water treatment models within SIMA-PRO have pre-determined emissions that result from water purification and the quality of the resulting effluent is assumed to be within legislative standards. Solid emissions were either reused as a fertilizing agent if it occurred in a ash form (Leibbrandt (2010)) or used to supplement the Gasification feed if it occurred as an oily sludge (Marano (2003)). Refer to APPENDIX 4.2

Data for emissions that was generated by the GFT gasoline in a standard passenger vehicle engine was determined from the GREET 2.7 model. The emission data generated by the GREET 2.7 model were calculated for fossil derived fuels which meant that it contained appreciable amounts of sulphur dioxide emissions. The SO_2 emission was removed to reflect the zero sulphur content of the gasoline resulting from the current process model for converting bagasse to FT transportation fuels, because this model contains a Rectisol unit that removes any trace amounts of sulphur present in the original biomass (Kreutz *et al.* (2008)). For the combustion of bioethanol, emission data for a passenger car engine that was modified for ethanol combustion was derived from Da-Silva *et al.* (2007). The ethanol used in the experiments of Da-Silva *et al.* (2007) was free of sulphur, but the ethanol resulting from the biological process in this study contained 50ppm of sulphur. Thus, the results from Da-Silva *et al.* (2007) was adjusted to reflect the amounts of sulphur dioxide that would result from the ethanol resulting from the models in this study. The engines exhaust characteristics resulting from gasoline and ethanol is presented in Table 15. The particulates were not measured by Da_Silva *et al.* (2007), so it was assumed that the engine would emit the same level of particulates as the gasoline engine.

Table 15: Engine Exhaust Data

Gasoline Engine Exhaust Emissions - Greet 2.7	grams/km
VOC: exhaust	0.0758
VOC: evaporation	0.0360
Carbon monoxide	2.3271
Nitrous Oxides	0.0876
Particaltes	0.027
N2O	0.0075
Carbon dioxide	234.1926
Ethanol Engine Emmisions - Da_Silva et eal 2007	grams/km
Carbon dioxide	246
Carbon monoxide	8.78
Hydrocarbons, unspecified	0.43
Sulfur dioxide	0.0065
Nitrous Oxides	0.79
Particaltes****	0.027

4.3.3.2 Allocation

The details of the allocation values that were used to distribute impacts between the products of each scenario are shown in Table 16. These are economic based allocations (given as percentages), for reasons described in Section 4.1.4.2. These values only apply to the product orientated LCAs and not the PEA's.

Table 16: Details of the Allocation of Impacts

	GASOLINE (%)	DIESEL (%)	ELECTRICITY (%)	ETHANOL (%)
GFT	34.894	52.202	12.904	
SSF			9.483	90.517
SHF			9.483	90.517

For the biological scenarios, the co-product was electricity where as the GFT scenario had electricity and diesel as co-products. Even though it may seem from Table 16 that diesel is the primary product of the GFT process, the gasoline is the FT fuel of importance here, since the study revolves around the comparison of ethanol and gasoline as products that deliver a gasoline driven vehicle for 1 km. Impacts were allocated between the co-products based on the revenue that each product produces. The prices on which the revenues were based for this particular purpose were the forecasted averages over the plant life. This was done to ensure that the results of the environmental impacts are statistically representative over the plants life. This kind of sensitivity that is induced by allocation methods could otherwise be handled through Monte Carlo simulation methods, but that kind of detail was not deemed to be part to be within the scope of a comparative exercise such as this.

4.3.4 Impact Category Assessments

4.3.4.1 Process Environmental Assessment in Locality

The results of the PEA of all the processes were firstly interpreted in terms of the local geographical situations (Burgess and Brennan 2001) in order to understand the absolute nature of environmental impacts that the relevant technologies impose. Thus, the cause for any excessive environmental burdens can be addressed for particular processes. These results are depicted in Figure 20.

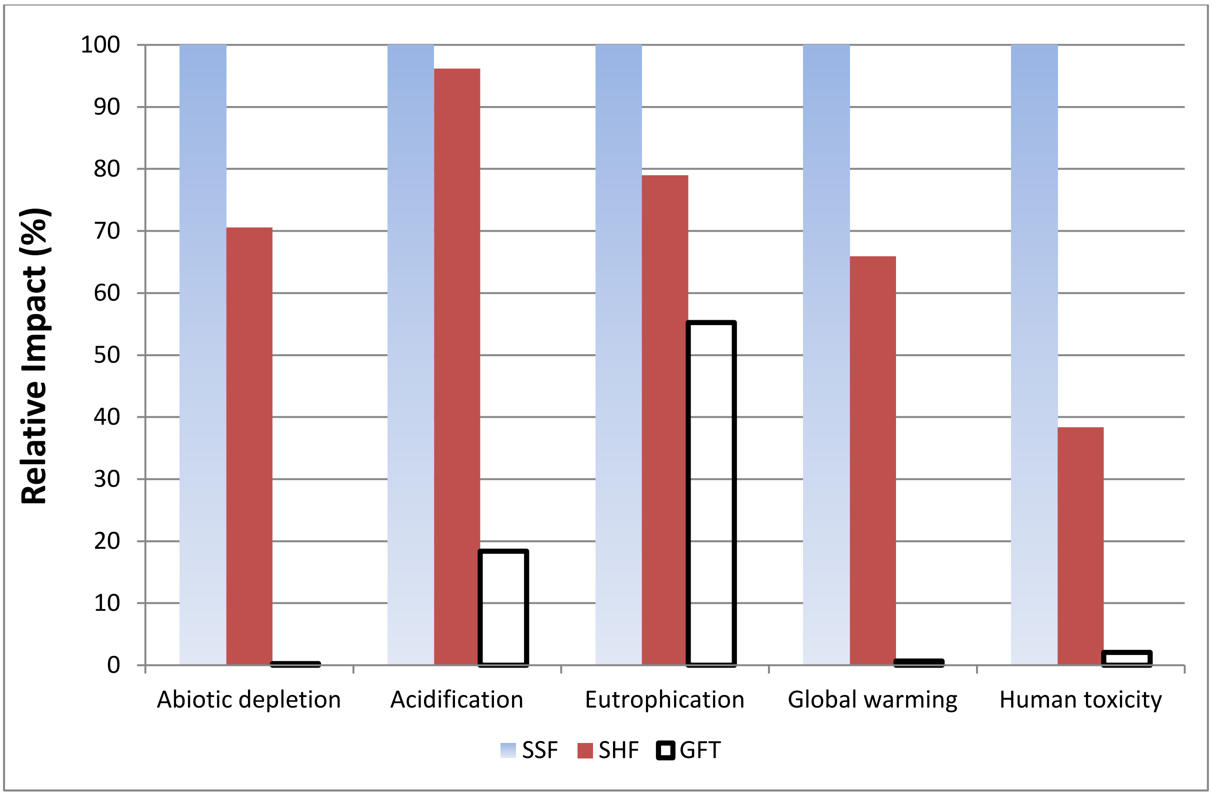

Figure 20: PEA Category Impacts

The geographical location of the processing facility was proposed in Leibbrandt (2010) to be in the Northern Natal/Swaziland region. These regions are generally described as tropical with high rainfall and vegetation consisting of afromontain forests. With such areas, acidification (Burgess and Brennan (2001)) would be of critical significance, while categories such as eutrophication of water bodies would be of less concern, since nutrients would likely flow to the ocean in a river with high volumetric flow due to coastal proximity of plant. The GFT process had the lowest potential for acidification, mostly because there wasn't any use of sulphur compounds in the process.

Human Toxicity is always a critical concern, even more so when the locality has a thick population density, and as depicted in Figure 20, the SSF process had the harshest impacts on human health. The absolute potentials measurements of Human Toxicity for the cell propagation schemes in the SHF and SSF was 171 and 522kg 1,4-DB eq. respectively, showing that the cell propagation scheme of the SSF is indeed the critical factor in this regard, due its intensity on nutrient demands (as by the recipe of Rudolf *et al.* (2007)). Furthermore, the cell propagation scheme in the SSF process relied entirely on an external glucose supply from a sugar refinery because a hydrolysate stream that is rich is glucose monosaccharides is not present in a SSF process configuration. If it is considered that the difference in ethanol conversion between the two process configurations is not that pronounced, then the nutrient intensity of the SSF cell propagation could be relaxed to diminish its impact on human health.

The SSF also had the highest potential for Global Warming, or for a more appropriate contextualisation for a geographical locality, Greenhouse Gas emissions. This occurred from the process demands of natural gas to supplement the excessive energy needs. Furthermore, SSF process had the highest energy demands for its downstream sections due to the excessive dilution in the fermentation-saccharification sections.

The GFT process had the lowest environmental impacts across most categories and only showed a significant contribution to eutrophication impacts. As stated above, eutrophication impacts would not be of critical concern in the current locality. Furthermore, this result is a reflection on the sheer quantity of reaction water that the Fischer-Tropsch reactor generated, and not of any significant amounts of phosphate that would enrich a receiving body of water with nutrients. Eutrophication of water bodies arise primary from the introduction of phosphates in industrial (and suburban) effluents. It is not expected that significant amounts of phosphates are found in the GFT scenario effluent because no phosphorous compounds were in use anywhere. With regards to the biological models, however, phosphates are required for cell growth. Thus, it would be expected that all phosphates be integrated into the fermenting organisms, and would end up in combustion products, rather than being any significant constituents of the water effluent.

4.3.4.2 **Product Life Cycle Assessments WRT Global Impact (Product LCA)**

Figure 21 depicts the environmental impacts associated with the amount of fuel product of each process that is required to deliver the functional unit for the entire life cycle. The impacts that are shown here can be considered as the global impacts of the systems, because each component or environmental flow of the life cycle that is shown in Figure 19 could potentially occur at various geographical locations. Thus, Figure 21 is not expected to correspond in any way to Figure 20 of the PEA because of the functional units in use and because the Product LCA includes all phases that life cycle entails. The most notable difference is that the SHF scenario, not the SSF scenario has now been rendered as the option which has posed the most severe environmental burdens. This is because the SSF scenario yielded ethanol at a greater production rate which implied that its processing impacts were diminished through a higher degree of scaling for the functional unit.

The impacts associated with the GFT scenario were the lowest even though the production rate of gasoline (which is required to deliver the functional unit) was lower than either of the ethanol production rates of the biological scenarios. This is because the allocation of impacts associated with the product stages prior to engine combustion are significantly less with the GFT gasoline in relation to the biological processing scenarios. Furthermore, the quantity of ethanol required for the functional unit is 1.54 times greater than that of gasoline, due to the higher HHV of gasoline.

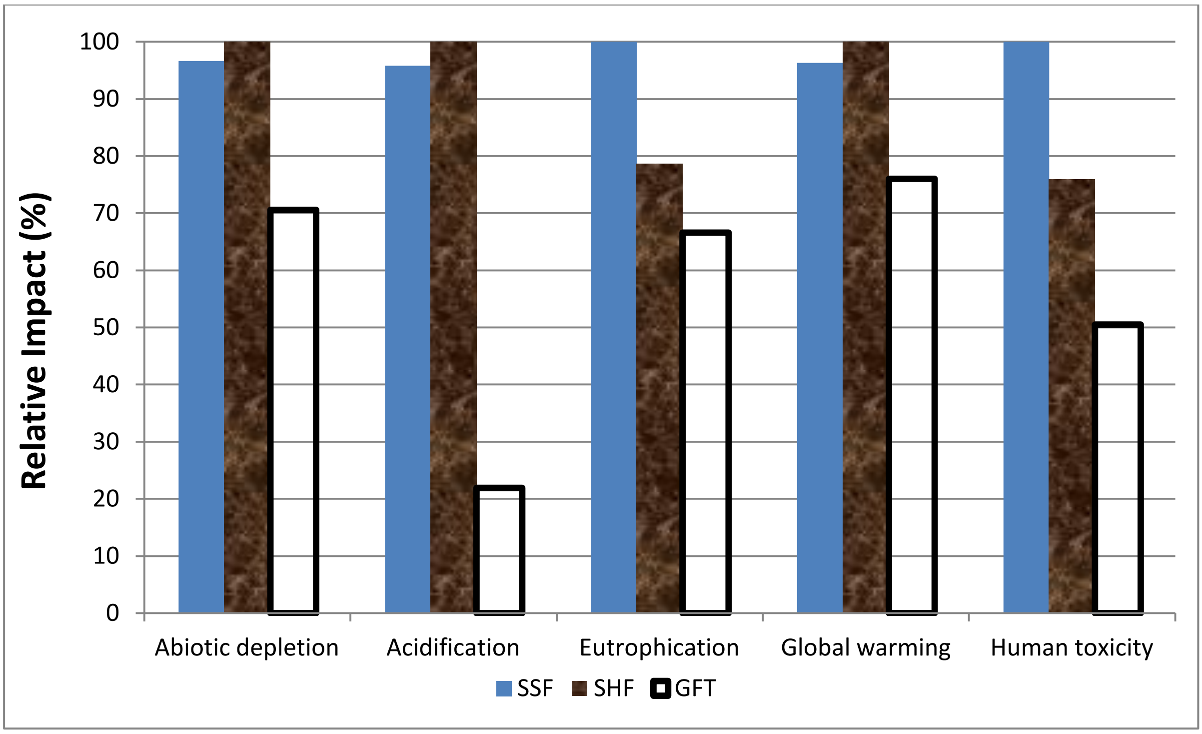

Figure 21: Product LCA Category Impacts

Comparing Figure 21 and Figure 20 also indicates that the product phases apart from the production phase can make significant contributions to the overall impact of a life cycle. For example, Figure 20 (PEA) shows that the human health impacts that were associated with the GFT production process were negligible, while from the perspective of a Product orientated LCA, the same scenario yielded significant impacts human health. This deduces that the toxicity emissions of the Transporting and Engine Combustion phases were greater than the production process itself.

4.3.4.3 Sensitivity of LCA results to Design Assumptions based on Technological Maturity

The absolute values of the impacts associated with the SHF process are not likely to improve significantly. This is because the data used to populate the Life Cycle Inventory was sourced from a simulation model that was based on experimental data and on flow-sheets that have been verified on a pilot scale (Aden *et al.* (2002)). Thus, it can be safely said that technology of the SHF scenario is on a pre-industrial level.

For the SSF scenario, conservative design assumptions were made with regards to the % solids content, because the technology has not officially been verified on a pilot scale. Due to the superior performances that SSF configurations have consistently shown on a laboratory scale over SHF, it can be said that SSF technology is only at a pre-pilot level. Thus, if pilot data can verify a high % solids level for SSF operations, the absolute impacts of Global Warming and Abiotic Depletion will decrease. This would also cause that the SSF displays the lesser impact for these categories in terms of the PEA.

Coal-to-liquid plants (CTL) have long been employed on a commercial scale while Biomass-to-liquid plants (BTL, i.e. the GFT plant) have not yet been implemented on a pilot scale. Since a large portion of the technology of the BTL plant is based on that of the CTL technology, then it can be said that the technology of the GFT scenario is commercially mature for those technologies. This level of maturity would not hold for the specific features of the BTL plant, such as the biomass handling and gasification. These technologies have mainly been tested on laboratory scale (Leibbrandt (2010)), and in isolation on a pilot scale (Kreutz *et al,* (2008)). As these technologies mature, it would not be expected that the environmental impact posed by the GFT scenario improves by any significant margin, because an improvement in the maturity is not expected to confirm a superior operating specification, as was shown to be necessary with the SSF scenario. Maturation would only confirm lower capital expenditure as technological maturity approaches the nth scenario.

4.4 CONCLUSIONS

The GFT scenario had the best environmental performance, irrespective of whether it was a LCA approach or a PEA approach. The energy efficiency of the GFT and the independence of the process to an excessive use of chemicals was shown to account for this occurrence.

The SSF scenario had the harshest impacted on human toxicity due to the cell propagation scheme in the production process. Even though this result conformed to the PEA, it is not necessarily a true reflection on the actual local reality because the nutrient supply could come from abroad. Nevertheless, the PEA results shows that impacts on Human Health of the SHF is 38% of the impact associated with the SSF. Given that the biomass output from the cell propagation scheme of the SSF is 7.5 times the mass of that of the SHF, and the overall conversion across the SSF is only 23% greater than the SHF (68% to 45%), then the cell propagation scheme of the SSF process could, at the least, be scaled down by a factor of 3. The overall SSF process should still be superior to the SHF, but with only +/-50% of the current potential for human health impacts, in the context of the PEA. This would also have produced economic advantages by reducing the expenditure on process chemicals and capital expenditure on the cell propagation scheme.

The assumption that saccharification-fermentation configuration in the SSF operated under more dilute conditions showed additional consequences in terms of environmental impacts, specifically the relatively high Abiotic depletion. From an energy perspective, the SSF scenario required more natural gas to supplement the energy which resulted in it having higher Greenhouse Gas Emissions. Thus, as it was stressed earlier, pilot plant data is needed to confirm a high slurry density for a SSF configuration. If it could be shown that the SSF configuration could indeed operate under thicker solids loadings, then the demand on Abiotic resources and the Greenhouse gas potential will diminish by up to 30%

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5 OVERALL CONCLUSIONS

5.1 GENERAL

Based on the overall view of the energy, economic and environmental comparisons of the three process scenarios, the following conclusions can be drawn:

- The GFT had the best energy efficiency, both in terms of liquid fuel production and overall energy efficiency. The high energy efficiency of the GFT process was further reflected by the potential of this process to generate the highest revenue and gross income, when the operating expenditure was accounted for. The pinch point retrofit on the GFT process model yielded a significant generation of excess electricity for exporting, to generate additional revenue which showed that pinch point retrofits contributed significantly to the energy efficiency. With regards to the biological models however, pinch point retrofits did not generate benefits that were as pronounced as it had for the GFT scenario, because the temperature of the hot steams, which is directly indicative of the heat quality was too low. The resulting benefits produced a successful economic outcome which otherwise would have been a failure without pinch point energy integration.

- The environmental burdens of the GFT scenario were also lower than the biological process across most categories of impacts, whether the approach taken to the LCA analysis was product orientated or process orientated. The relatively low Abiotic Depletion and Global Warming potentials that are associated with the GFT scenario were also indicative of the optimal energy efficiency.

- The GFT scenario was likely to be the worst option for investments because Monte Carlo Simulations that were carried out on the Key Economic Indicators revealed that their likelihood of exceeding the stipulated limits were too high. The primary contributing fact to this result lies in the enormity of the capital investment. Thus, the cash returns were not able to produce reasonable payback on this investment even though it was higher than that of any of the biological models. Monte Carlo Simulations also showed that this investment will not even be feasible for the low IRR expected from the IDC, even when the deterministic model had shown that it was feasible.

- The liquid fuel and overall energy efficiency of the SSF process were marginally higher than that of the SHF process. As a result, there was better potential for a net cash income, which ultimately resulted in a financial performance that was similar in terms of the NPV and IRR, despite the excessive expenditure on chemicals for cell propagation. The additional energy efficiency, however, arose at excessive environmental impacts on Human Toxicity, as indicated by the Process Orientated LCA. Furthermore, it was observed that the basic flow sheet for the biological models was not an optimum design for an SSF configuration, because the recycling scheme only favoured an SHF type plant. There is further potential for this scenario if more attention is given to the actual flow-sheet. For example, the residence time in the SSF reactor could be extended, or if a recycling system could be tailor made to suit the SSF flow-sheet.

- Energy demands of the SSF process were higher than that of the SHF, because of the assumption that the configuration of the SSF process operated at dilute conditions. As mentioned earlier, this assumption arose from the unavailability of pilot plant data to confirm that the continuous SSF configuration can operate at the same solids loadings as the SHF. The consequences of this assumption also affected the LCA analysis, because the SSF scenario had an exaggerated affect on green house gas emissions and Abiotic Depletion.

- Both the SSF and SHF scenarios displayed complete investment feasibility when intrinsic risks and uncertainties associated with market prices are accounted for with Monte Carlo economics. The difference between the two were marginal because the lower operating costs of the SHF played off against the higher ethanol production rates of the SSF scenario. The higher ethanol production of the SSF scenario caused that it had a better Product Orientated LCA performance, but the non-optimised operating conditions caused that the process had the most severe effects on the local environment, as shown by the PEA.

- The technological endpoint of biological processing is Consolidated Bioprocessing (CBP). Considering that this technology is basically the SSF technology without enzyme addition, the expected IRR of such technology would be ± 26% by estimation. With regards to the LCA of this futuristic scenario, it would be represented by that of the SSF without the impacts associated with enzyme production.

5.2 HYPOTHESIS TESTING

With reference to the general conclusions drawn above, the statuses regarding the Hypothesis of overall project (see Section 1.3) can be established.

1.1 Accepted, because capital costs were reduced for the SSF and because the energy efficiency improved, leading to better liquid fuel outputs.
1.2 Accepted as is. Furthermore, heat integration was to be maximised for the optimum effect.
1.3 Accepted, because the overall energy of the GFT scenario was about 21% better than the SSF scenario. The pinch point retrofit contributed significantly to this vast difference due to the benefits that it yielded for the GFT scenario.
1.4 Accepted as is.
1.5 Rejected, because the GFT environmental impacts were considerably lower than that any of the biological scenarios, in the contexts of both product and process orientated approaches.

6 RECOMMENDATIONS

Based on the general conclusions, the following are recommended:

6.1 IMPROVING THE SSF EFFICIENCY

A number of items regarding in the SSF scenario were identified as areas to be considered for further investigation, so that the energy efficiency can improve to rival that of the GFT scenarios:

1. To confirm that the SSF can operate under high solids concentration, so that energy demands for distillation and evaporation can be reduced to that of the SHF scenario. This would improve the sales of electricity by 0.36USD/MW_{HHV} of energy input.
2. To extend the residence time of the SSF reactor to maximise the hydrolysis of glucose, and its subsequent conversion into ethanol, or to redesign the recycling system so that un-hydrolysed cellulose can be given a second pass, so that the overall conversion of biomass to ethanol can improve. This could improve the energy efficiency from 42.8% to 46% (refer to section 2.6.4.1).

These recommendations will improve the overall profitability of the SSF, and reduce the SSF dependency on Abiotic Resources and its potential for Greenhouse Gas Emissions by 30%.

6.2 LESSENING THE SSF IMPACT ON HUMAN HEALTH

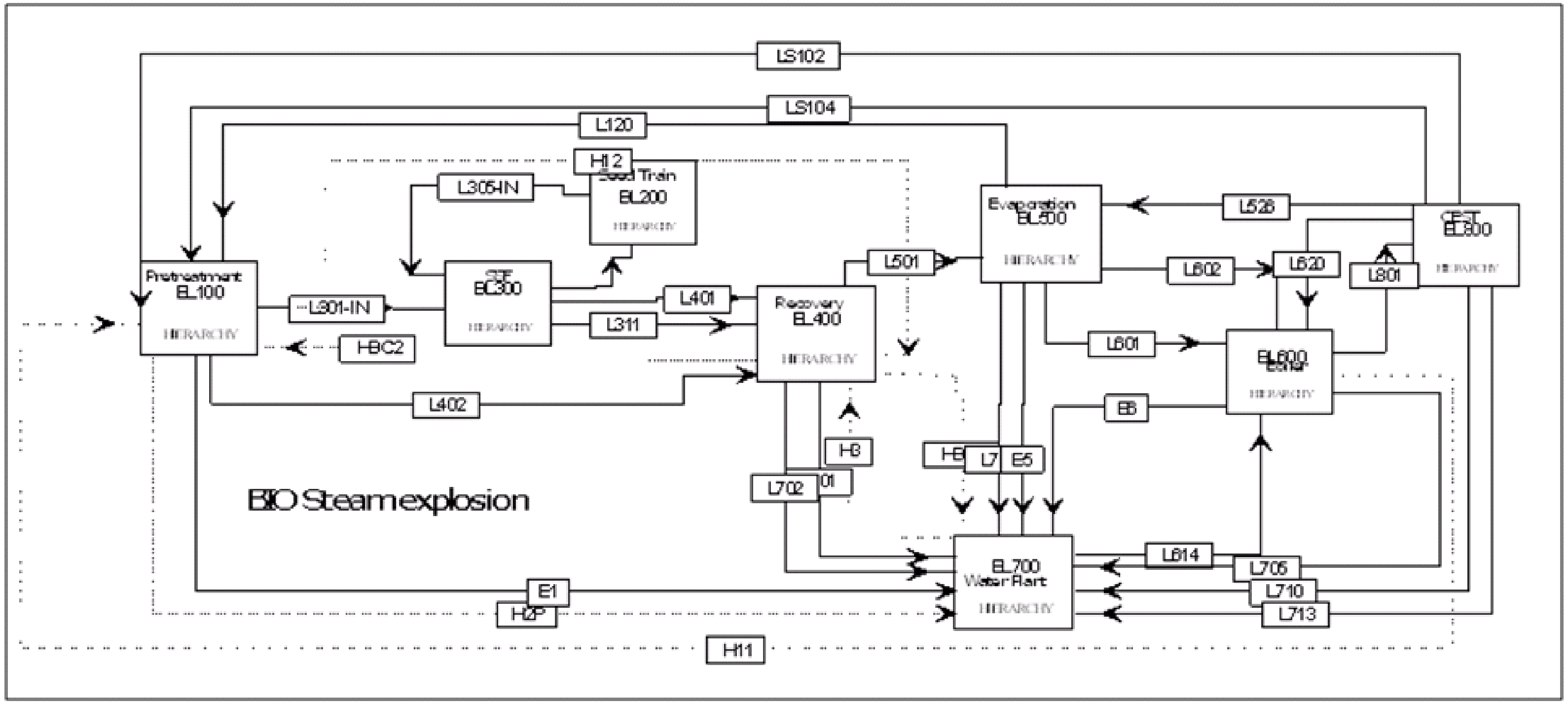
The SSF scenario was very nutrient and chemical intensive due the aggressive manner the yeast cells were propagated. Considering that the SHF scenario had similar ethanol conversions at much lower impacts, the human health impacts of the SSF process can certainly be reduced by employing a cell propagation recipe that is less intensive. Thus, it is recommended that such a recipe be found in experimental literature as a “friendlier” alternative to the current one employed by the SSF, which can reduce the impact on human health of the PEA by about 50%.

6.3 IMPROVING THE PROFITABILITY OF THE GFT SCENARIO

The GFT scenario is currently not attractive option from an economic perspective. To improve its economic appeal, it would be recommended to investigate the extent to which the profitability increases by improving electricity production efficiency by integrating a gas turbine with steam turbines.

APPENDIX 1: MASS BALANCES AND PFD'S

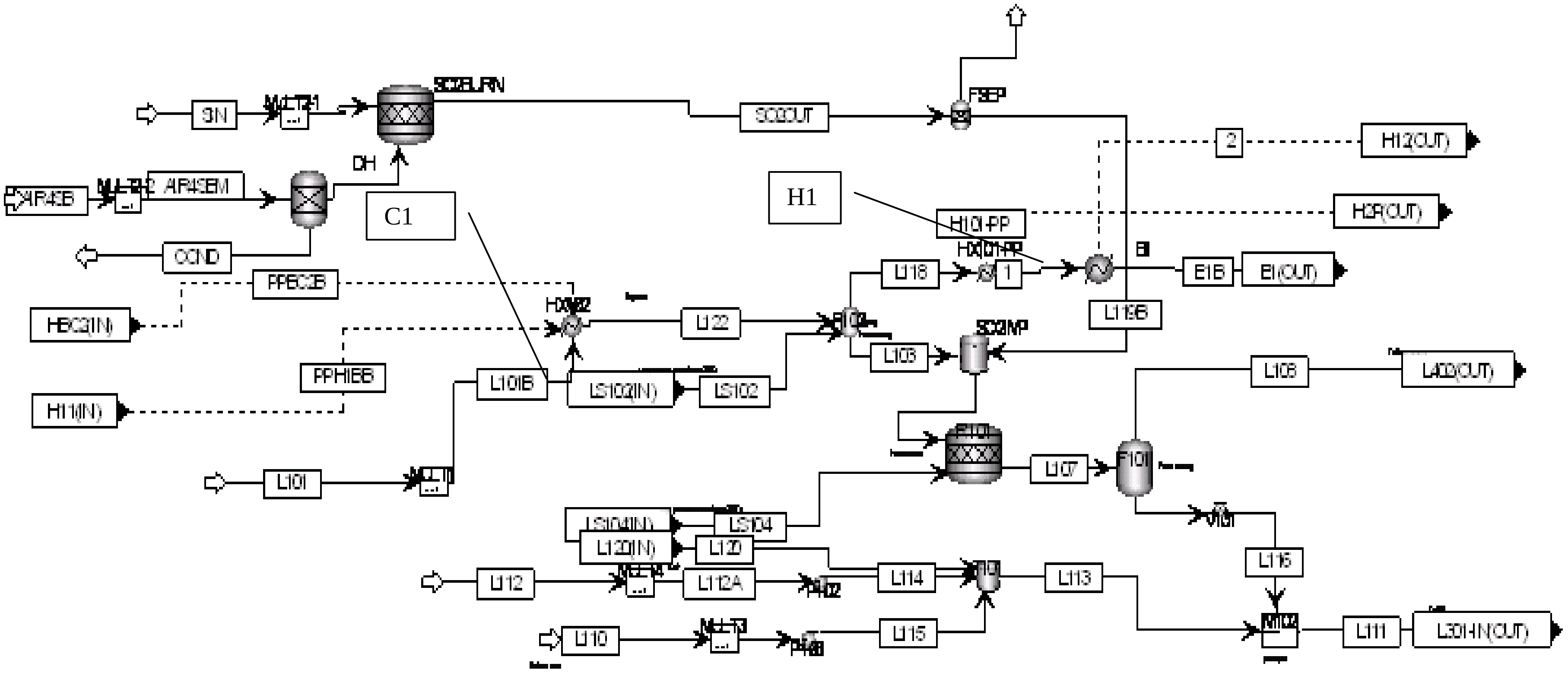
Generic BFD (Block Flow Diagram) reiterated for Biological Scenarios.



APPENDIX 1.1: PROCESS FLOW DIAGRAMS OF THE BIOETHANOL SCENARIOS

These PFDs will serve as generic for both the SHF and SSF since they are identical in most respects, besides AREA 100, 200 and 300.

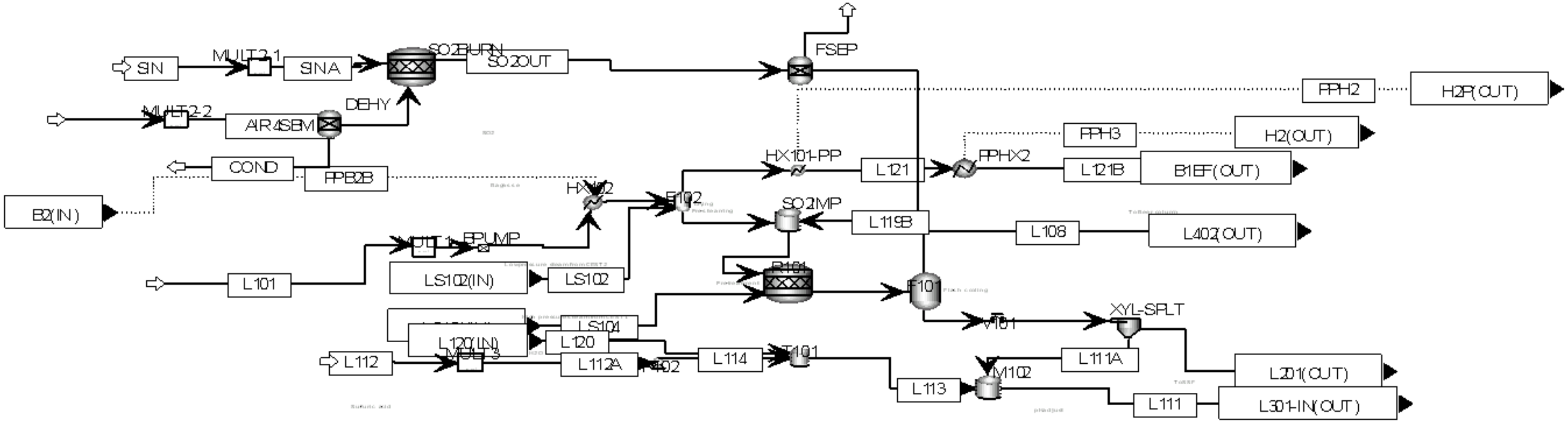
AREA 100 SHF


SIN
MULT21
SO2BURN
SO2OUT
FSEP
AIR4OB
MULT12
AIR4SEM
DH
C1
OCND
PPEC2B
HEX2(IN)
HX102
PPHIBB
H11(IN)
L101
MULT1
L101B
LS102(IN)
LS102
L122
H1
H101-PP
HX101-PP
1
2
H12(OUT)
H2R(OUT)
EII
E1B
EI(OUT)
L118
L119B
L103
SO2MP
L108
LA02(OUT)
L107
F101
V101
LS104(IN)
LS104
L120(IN)
L120
L112
MULT4
L112A
L114
L113
L116
L110
MULT3
L115
M102
L111
L301HN(OUT)

AREA 100 SHF

	1	AIR4SB	AIR4SBM	C	COND	DA	E1B	F1	L101	L101B	L103	L107	L108	L109	L110	L111	L112	L112A	L113	L114	L115	L116	L118	L119B	L120	L122	LS102	LS104	PTF	SIN	SINA	SO2OUT
Temperat	375	298.1	298.1	298.1	298.1	298.1	373.1	493.2	298.1	298.1	420	483.1	373.7	373.7	298.1	338.1	298.1	298.1	298.2	298.2	298.2	373.7	420	493.2	298.1	373.3	598.3	733	372.4	298.1	298.1	493.1
Pressure	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	4.336	19.5	1	1	1.013	1.013	1.013	1.013	1.013	1.115	1.115	1	4.336	1.013	1.013	1.013	4.458	13	1.013	1.013	1.013	1.013
Mass VFra	1	1	1	0	0	1	0.552	1	0	0	0	0	1	0	0	0	0	0	0	0	0	0	1	1	0	0.007	1	1	0.06	0	0	1
Mass SFra	0	0	0	0	0	0	0	0	0.463	0.463	0.459	0.333	0	0.385	0	0.133	0	0	0.002	0	0	0.385	0	0	0.007	0.463	0	0	0.455	0	0	0
Mass Flow	9449.907	65.384	8035.798	0.587	104.465	7931.333	9449.907	6784.734	2000	245803.2	247475.4	250884.3	33553.29	217331	0.005	632378.1	2510	308483	415047.1	308483	0.587	217331	9449.907	2295.606	106563.5	245803.2	11122.07	1113.303	249771	9.349	1149.007	9080.34
Mass Flow	9449.907	65.384	8035.798	0.587	104.465	7931.333	9449.907	6784.734	2000	245803.2	247475.4	250884.3	33553.29	217331	0.005	632378.1	2510	308483	415047.1	308483	0.587	217331	9449.907	2295.606	106563.5	245803.2	11122.07	1113.303	249771	9.349	1149.007	9080.34
Volume Fl	266776.6	930.934	114413.3	0.005	1.746	112065.1	146584.8	161328.5	27.56	3387.174	3625.441	3927.724	904043.4	3026.291	0	10001.29	41.952	5155.986	6912.3	5155.963	0.005	3026.291	68726.67	24090.71	1756.312	53499.91	113268.5	4737.618	388020.6	0.386	47.411	185447.8
Enthalpy	-29.965	-0.003	-0.336	-0.001	-0.396	0	-32.262	0.327	-5.177	-636.244	-624.466	-625.248	-101.4	-523.847	0	-2072.42	-9.518	-1169.72	-1560.89	-1169.72	-0.001	-523.847	-29.791	-2.47	-391.164	-624.459	-34.013	-3.333	-627.018	0.019	2.287	-2.143
Mass Flow																																
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	3858.771	0	3858.771	0	4171.68	0	0	312.909	0	0	3858.771	0	0	312.909	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	406	49898.05	49898.05	42952.24	0	42952.24	0	42988.78	0	0	36.54	0	0	42952.24	0	0	36.54	49898.05	0	0	49898.05	0	0	0
XYLOSE	0	0	0	0	0	0	0	0	0	0	0	22597.05	0	22597.05	0	25509.97	0	0	2912.92	0	0	22597.05	0	0	2912.92	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	200	24580.32	24580.32	3932.851	0	3932.851	0	3952.614	0	0	19.763	0	0	3932.851	0	0	19.763	24580.32	0	0	24580.32	0	0	0
LIGNIN	0	0	0	0	0	0	0	0	255	31339.91	31339.91	31339.91	0	31339.91	0	31497.4	0	0	157.487	0	0	31339.91	0	0	157.487	31339.91	0	0	31339.91	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	49.057	0	0	49.057	0	0	0	0	0	49.057	0	0	0	0	0	0	0
MICROOR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	497.855	0	0	497.855	0	0	0	0	0	497.855	0	0	0	0	0	0	0
UNKNOW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINO	0	0	0	0	0	0	0	0	0	0	0	1842.4	0	1842.4	0	2377.29	0	0	534.89	0	0	1842.4	0	0	534.89	0	0	0	0	0	0	0
GALACTO	0	0	0	0	0	0	0	0	0	0	0	778.921	0	778.921	0	1005.059	0	0	226.138	0	0	778.921	0	0	226.138	0	0	0	0	0	0	0
MANNOS	0	0	0	0	0	0	0	0	0	0	0	194.73	0	194.73	0	251.265	0	0	56.535	0	0	194.73	0	0	56.535	0	0	0	0	0	0	0
ARABINA	0	0	0	0	0	0	0	0	17	2089.327	2089.327	468.009	0	468.009	0	470.361	0	0	2.352	0	0	468.009	0	0	2.352	2089.327	0	0	2089.327	0	0	0
MANNAN	0	0	0	0	0	0	0	0	2	245.803	245.803	70.546	0	70.546	0	70.9	0	0	0.355	0	0	70.546	0	0	0.355	245.803	0	0	245.803	0	0	0
GALACTA	0	0	0	0	0	0	0	0	8	983.213	983.213	282.182	0	282.182	0	283.6	0	0	1.418	0	0	282.182	0	0	1.418	983.213	0	0	983.213	0	0	0
CELLOB	0	0	0	0	0	0	0	0	0	0	0	316.02	0	316.02	0	316.02	0	0	0	0	0	316.02	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	75.057	0	0	75.057	0	0	0	0	0	75.057	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	0	0	0	0	75	9217.62	9217.62	9217.62	0	9217.62	0	9270.565	0	0	52.945	0	0	9217.62	0	0	52.945	9217.62	0	0	9217.62	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	37	4547.359	4547.359	4547.359	0	4547.359	0	4570.21	0	0	22.851	0	0	4547.359	0	0	22.851	4547.359	0	0	4547.359	0	0	0
ETHANOL	0.201	0	0	0	0	0	0.201	0	0	0	0.327	0.379	0.29	0.089	0	5.639	0	0	5.55	0	0	0.089	0.201	0	5.55	0	0.528	0.053	0.327	0	0	0
H2O	9432.873	0.85	104.465	0	104.465	0	9432.873	0	1000	122901.6	124513.4	122041.6	30924.86	91116.79	0	500278.2	2510	308483	409161.4	308483	0	91116.79	9432.873	0	100678.6	122901.6	11044.67	1105.556	124513.4	0	0	0
N2	0	49.6	6095.957	0	0	6095.957	0	6095.957	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	6095.957
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	14.934	1835.376	0	0	1835.376	0	688.777	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	688.777
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	1.336	0	0	0	0	0	1.336	0	0	0	32.325	3561.813	565.448	2996.364	0	3766.333	0	0	769.969	0	0	2996.364	1.336	0	769.969	0	33.661	3.369	32.325	0	0	0
NH4SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	15.249	0	0	0	0	0	15.249	0	0	0	21.366	2320.637	2037.185	283.452	0	283.452	0	0	0	0	0	283.452	15.249	2295.606	0	0	36.615	3.665	2316.972	0	0	2295.606
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0.004	0	0.001	0	0.292	0	0	0.293	0	0	0.001	0	0	0.066	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0.003	0	0.001	0	0	0	0	0	0	0	0.001	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	0.02	0	0	0	0	0	0.02	0	0	0	1.605	555.949	24.582	531.366	0	681.432	0	0	150.066	0	0	531.366	0.02	0	150.066	0	1.625	0.163	1.605	0	0	0
SO4--	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0.732	0	0	0.738	0	0	0	0	0	0.166	0	0	0	0	0	0	0
SULFU-01	0	0	0	0.587	0	0	0	0	0	0	0	0	0	0	0.005	0	0	0	0	0	0.587	0	0	0	0	0	0	0	0	0	0	0
HYDRAZI	0.228	0	0	0	0	0	0.228	0	0	0	4.743	5.241	0.925	4.316	0	4.316	0	0	0	0	0	4.316	0.228	0	0	0	4.971	0.498	4.743	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	9.349	1149.007	0

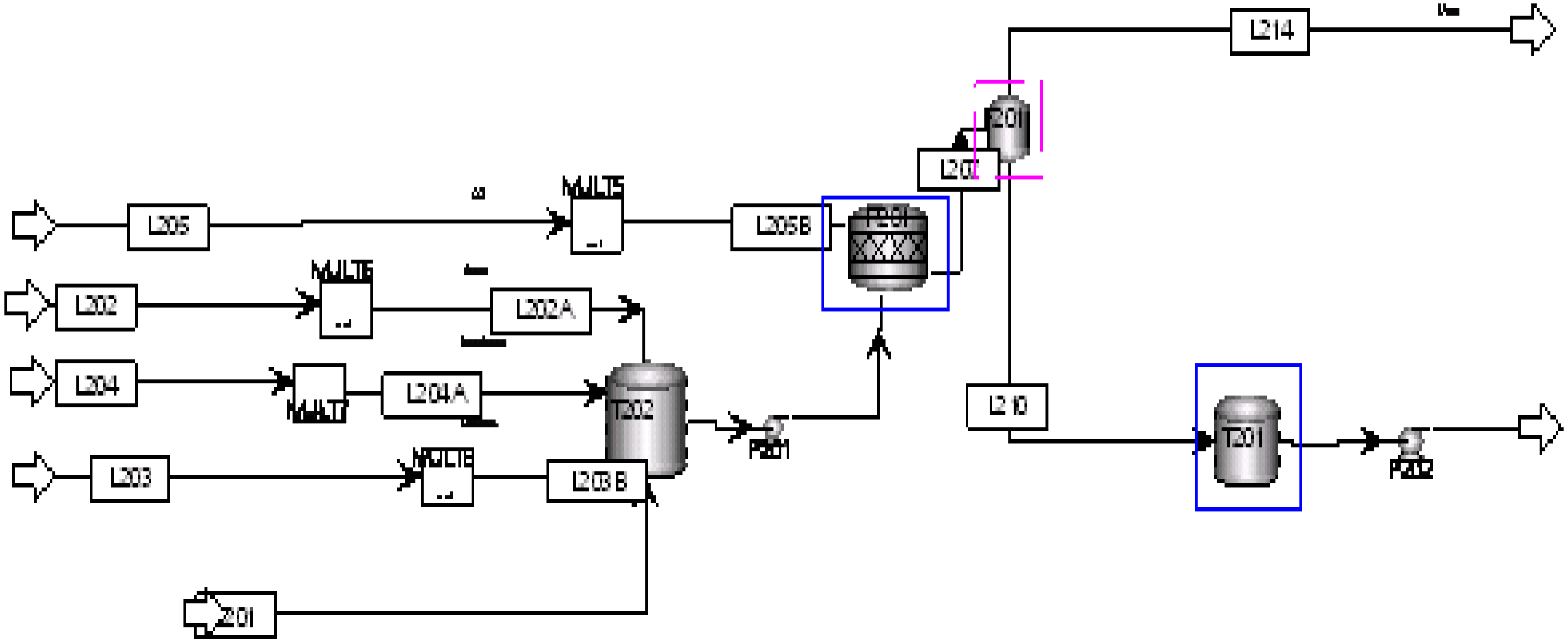
AREA 100 SSF


SIN
MULT2-1
SINA
SO2BURN
SO2OUT
FSEP
MULT2-2
AIR4SBM
DEHY
COND
FPB2B
B2(IN)
Bagasse
HX102
F102
HX101-PP
L121
PPHX2
L121B
B1EF(OUT)
PPH2
H2P(OUT)
PPH3
H2(OUT)
SO2MP
L119B
L108
L402(OUT)
To Beer column
L101
MULT1
BPUMP
LS102(IN)
LS102
R101
Pretreatment
F101
Flash cooling
V101
XYL-SPLT
LS104
L120(IN)
L120
L112
MULT3
L112A
P102
L114
T101
L113
M102
L111A
L201(OUT)
L111
L301-IN(OUT)
Sulfuric acid
pHadjust

AREA 100 SSF

	AIR4SB	AIR4SBM	COND	DA	L101	L101A	L101B	L103	L107	L108	L109	L111	L111A	L111B	L112	L112A	L113	L114	L116	L118	L119B	L120	L121	L121B	L122	LS102	LS104	O2-N2P	PTF	SIN	SINA	SO2OUT
Temperature K	298	298	298	298	298	298	298	420	483	374	374	312	374	374	298	298	298	298	374	420	493	298	375	373	373	571	701	493	372	298	298	493
Pressure bar	1	1	1	1	1	4	1	4	20	1	1	1	1	1	1	1	1	1	1	4	1	1	1	1	1	5	13	1	1	1	1	1
Mass VFrac	1	1	0	1	0	0	0	0	0	1	0	0	0	0	0	0	0	0	0	1	1	0	1	1	0	1	1	1	0	0	0	1
Mass SFrac	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr	65	8036	104	7931	2000	245803	245803	247987	251451	33674	217777	666468	206613	11164	2789	342773	459855	342773	217777	9489	2296	117083	9489	9489	245803	11673	1168	6785	250282	9	1149	9080
Mass Flow kg/hr	65	8036	104	7931	2000	245803	245803	247987	251451	33674	217777	666468	206613	11164	2789	342773	459855	342773	217777	9489	2296	117083	9489	9489	245803	11673	1168	6785	250282	9	1149	9080
Volume Flow l/min	931	114413	2	112075	28	3387	3387	3635	3939	907468	3034	10485	2863	171	47	5729	7667	5729	3034	69004	24091	1938	267864	158902	65985	111505	4681	161328	389300	0	47	185448
Enthalpy Gcal/hr	0	0	0	0	-5	-636	-636	-626	-627	-102	-526	-2227	-493	-32	-11	-1300	-1734	-1300	-526	-30	-2	-434	-30	-32	-624	-36	-4	0	-629	0	2	-2
Mass Flow kg/hr																																
GLUCOSE	0	0	0	0	0	0	0	0	3859	0	3859	4325	3538	321	0	0	787	0	3859	0	0	787	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	406	49898	49898	49898	42952	0	42952	43007	42952	0	0	0	55	0	42952	0	0	55	0	0	49898	0	0	0	49898	0	0	0
XYLOSE	0	0	0	0	0	0	0	0	22597	0	22597	21687	20717	1880	0	0	970	0	22597	0	0	970	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	200	24580	24580	24580	3933	0	3933	3953	3933	0	0	0	20	0	3933	0	0	20	0	0	24580	0	0	0	24580	0	0	0
LIGNIN	0	0	0	0	255	31340	31340	31340	31340	0	31340	31497	31340	0	0	0	157	0	31340	0	0	157	0	0	31340	0	0	0	31340	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	330	0	0	0	0	330	0	0	0	0	330	0	0	0	0	0	0	0	0	0	0
MICROORG	0	0	0	0	0	0	0	0	0	0	0	156	0	0	0	0	156	0	0	0	0	156	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINOS	0	0	0	0	0	0	0	0	1842	0	1842	2224	1689	153	0	0	535	0	1842	0	0	535	0	0	0	0	0	0	0	0	0	0
GALACTOS	0	0	0	0	0	0	0	0	779	0	779	940	714	65	0	0	226	0	779	0	0	226	0	0	0	0	0	0	0	0	0	0
MANNOSE	0	0	0	0	0	0	0	0	195	0	195	235	179	16	0	0	57	0	195	0	0	57	0	0	0	0	0	0	0	0	0	0
ARABINAN	0	0	0	0	17	2089	2089	2089	468	0	468	470	468	0	0	0	2	0	468	0	0	2	0	0	2089	0	0	0	2089	0	0	0
MANNAN	0	0	0	0	2	246	246	246	71	0	71	71	71	0	0	0	0	0	71	0	0	0	0	0	246	0	0	0	246	0	0	0
GALACTAN	0	0	0	0	8	983	983	983	282	0	282	284	282	0	0	0	1	0	282	0	0	1	0	0	983	0	0	0	983	0	0	0
CELLOB	0	0	0	0	0	0	0	0	316	0	316	290	290	26	0	0	0	0	316	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	0	0	0	0	0	0	0	0	171	0	0	0	0	171	0	0	0	0	171	0	0	0	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	75	9218	9218	9218	9218	0	9218	8497	8451	767	0	0	46	0	9218	0	0	46	0	0	9218	0	0	0	9218	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	37	4547	4547	4547	4547	0	4547	4570	4547	0	0	0	23	0	4547	0	0	23	0	0	4547	0	0	0	4547	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	7	0	0	0	0	7	0	0	0	0	7	0	0	0	1	0	0	0	0	0	0
H2O	1	104	104	0	1000	122902	122902	125025	122608	31045	91563	539333	83945	7618	2789	342773	455388	342773	91563	9471	0	112615	9471	9471	122902	11594	1161	0	125025	0	0	0
N2	50	6096	0	6096	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	6096	0	0	0	6096
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	15	1835	0	1835	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	689	0	0	0	689
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	0	0	0	0	0	0	0	32	3561	565	2996	3520	2747	249	0	0	773	0	2996	1	0	773	1	1	0	33	3	0	32	0	0	0
NH4SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCEROL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIENT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	22	2322	2038	284	260	260	24	0	0	0	0	284	16	2296	0	16	16	0	38	4	0	2318	0	0	2296
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	0	0	0	0	0	0	0	2	556	25	531	637	487	44	0	0	150	0	531	0	0	150	0	0	0	2	0	0	2	0	0	0
CA(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULAS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MGSO4-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HYDRAZIN	0	0	0	0	0	0	0	5	5	1	4	4	4	0	0	0	0	0	4	0	0	0	0	0	0	5	1	0	5	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	9	1149	0
SO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-02	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DIAMM-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ORTHO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

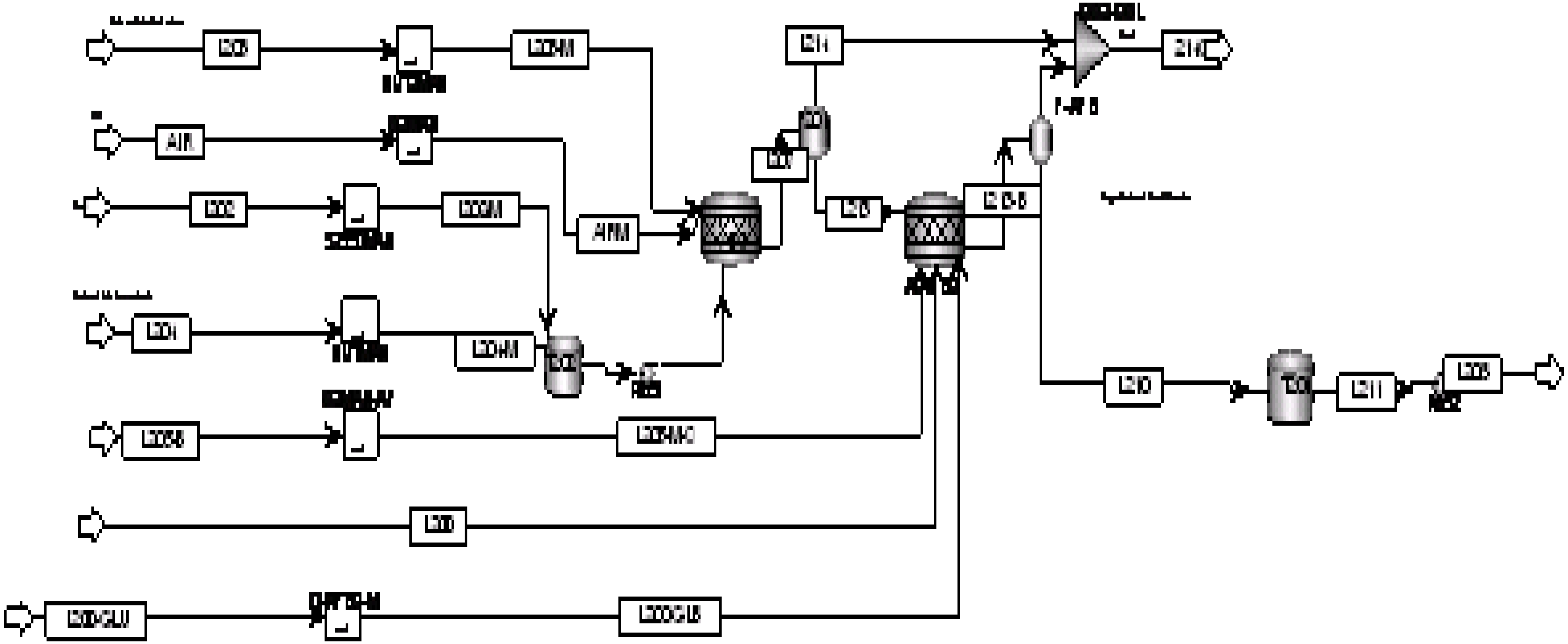
AREA 200 SHF



AREA 200 SHF

	L201	L202	L202A	L203	L203B	L204	L204A	L205	L205B	L206	L207	L208	L209	L210	L211	L212	L213	L214
Temperature	338	303	303	303	303	303	303	303	303	303	303	303	339	303	303	339	303	303
Pressure ba	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
Mass VFrac	0	0	0	0	0	0	0	1	1	0	0	0	0	0	0	0	0	1
Mass SFrac	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow k	63380	0	0	4	492	2	283	35	4302	60081	68456	3490	64155	60081	60081	64155	63571	4885
Mass Flow k	63380	0	0	4	492	2	283	35	4302	60081	68456	3490	64155	60081	60081	64155	63571	4885
Volume Flow	1009	0	0	0	9	25	3038	453	55693	940	58937	61	1020	940	940	1020	993	57945
Enthalpy Gc	-209	0	0	0	-1	0	0	0	0	-206	-216	0	-210	-206	-206	-210	-212	-4
Mass Flow k																		
GLUCOSE	4415	0	0	4	492	0	0	0	0	0	3336	3336	4907	0	0	4907	3336	0
CELLULOS	731	0	0	0	0	0	0	0	0	731	731	0	731	731	731	731	731	0
XYLOSE	2551	0	0	0	0	0	0	0	0	2551	2551	0	2551	2551	2551	2551	2551	0
XYLAN	395	0	0	0	0	0	0	0	0	395	395	0	395	395	395	395	395	0
LIGNIN	3150	0	0	0	0	0	0	0	0	3150	3150	0	3150	3150	3150	3150	3150	0
BIOMASS	5	0	0	0	0	0	0	0	0	348	348	0	5	348	348	5	348	0
MICROORG	50	0	0	0	0	0	0	0	0	50	50	0	50	50	50	50	50	0
UNKNOWN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINOS	238	0	0	0	0	0	0	0	0	238	238	0	238	238	238	238	238	0
GALACTOS	101	0	0	0	0	0	0	0	0	101	101	0	101	101	101	101	101	0
MANNOSE	25	0	0	0	0	0	0	0	0	25	25	0	25	25	25	25	25	0
ARABINAN	47	0	0	0	0	0	0	0	0	47	47	0	47	47	47	47	47	0
MANNAN	7	0	0	0	0	0	0	0	0	7	7	0	7	7	7	7	7	0
GALACTAN	28	0	0	0	0	0	0	0	0	28	28	0	28	28	28	28	28	0
CELLOB	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	8	0	0	0	0	0	0	0	0	8	8	0	8	8	8	8	8	0
EXTRACT	927	0	0	0	0	0	0	0	0	927	927	0	927	927	927	927	927	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	457	0	0	0	0	0	0	0	0	457	457	0	457	457	457	457	457	0
ETHANOL	1	0	0	0	0	0	0	0	0	1	1	0	1	1	1	1	1	0
H2O	49630	0	0	0	0	0	0	0	0	50261	50366	0	49630	50261	50261	49630	50261	105
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	19	1798	0	0	19	19	0	19	1779
O2	0	0	0	0	0	0	0	35	4302	1	2994	0	0	1	1	0	1	2993
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	1	127	0	0	124	127	0	127	124	124	127	124	3
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	377	0	0	0	0	0	0	0	0	376	377	0	377	376	376	377	376	0
NH4SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCEROL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIENT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	28	0	0	0	0	0	0	0	0	23	28	0	28	23	23	28	23	5
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	68	0	0	0	0	0	0	0	0	68	68	0	68	68	68	68	68	0
SODIU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULAS	143	0	0	0	0	0	0	0	0	143	143	0	143	143	143	143	143	0
FURFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	1	85	0	0	1	85	84	85	1	1	85	85	0
MGSO4-01	0	0	0	0	0	1	71	0	0	1	71	70	71	1	1	71	71	0
HYDRAZIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

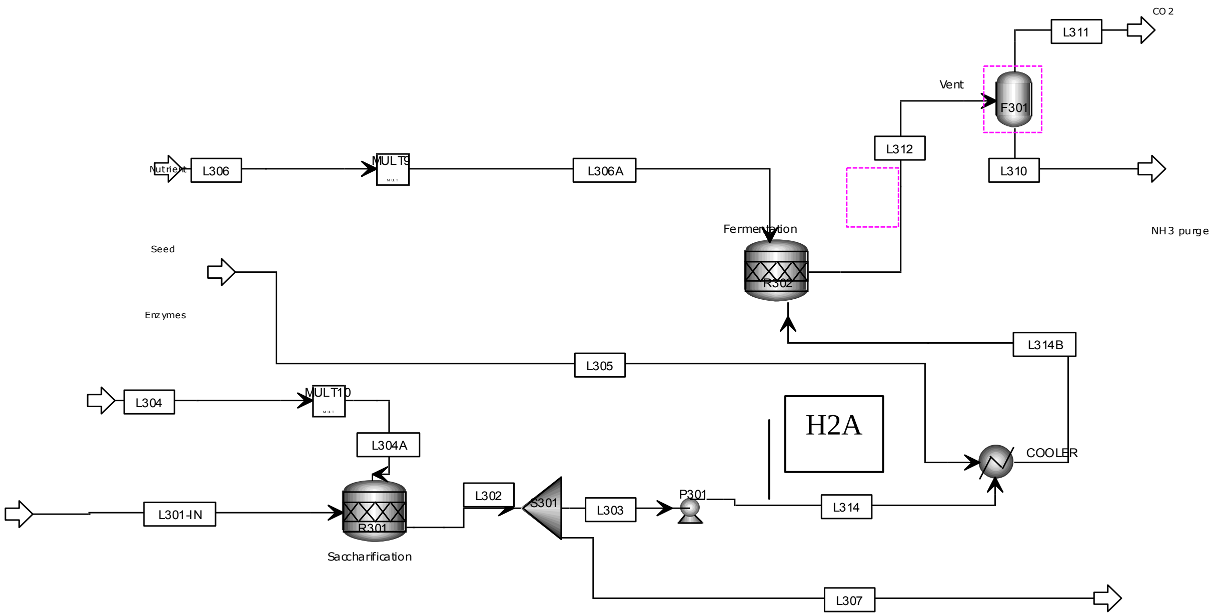
AREA 200 SSF



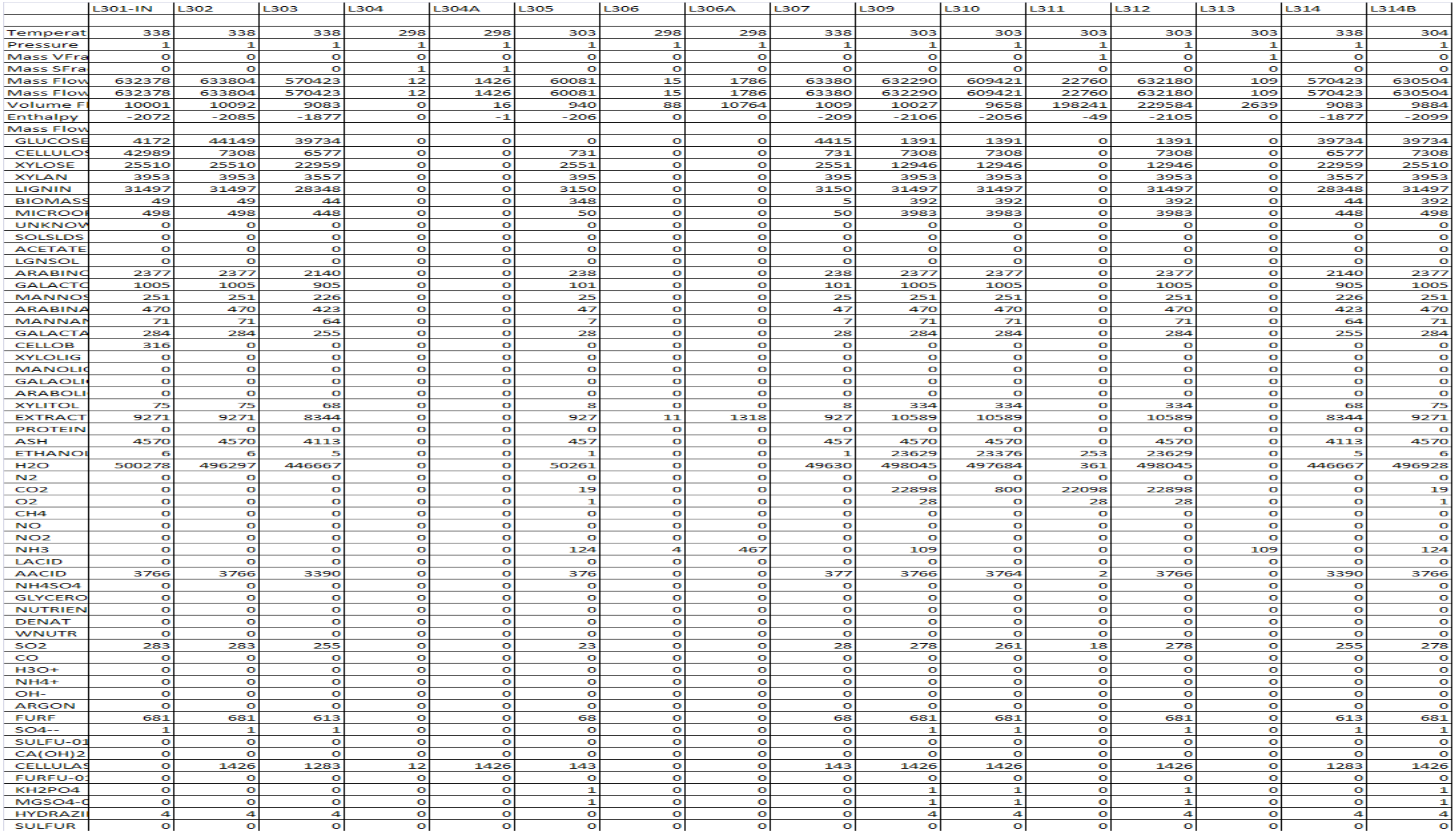
AREA 200 SSF

	214C	AIR	AIRM	L200	L200-GLB	L200-GLU	L202	L202-M	L203	L203-M	L204	L204-M	L205-B	L205-M-C	L206	L207	L208	L209	L210	L211	L212	L213	L213-B	L213C	L214	L214B
Temperat	303	298	298	374	298	298	303	303	303	303	303	303	298	298	303	303	303	303	303	303	303	303	303	303	303	303
Pressure	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
Mass VFra	1	1	1	0	0	0	0	0	0	0	0	0	1	1	0	0	0	0	0	0	0	0	0	0	1	1
Mass SFra	0	0	0	0	0	0	1	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	16103	158	734	11164	2308	494	148	690	1005	6121	990	2584	3000	14003	21400	10129	102	3274	21400	21400	3274	9330	36807	21502	798	15305
Mass Flow	16103	158	734	11164	2308	494	148	690	1005	6121	990	2584	3000	14003	21400	10129	102	3274	21400	21400	3274	9330	36807	21502	798	15305
Volume F	221247	2236	10392	171	41	9	2	8	18	109	17	44	42450	198146	332	11166	2	52	332	332	52	159	210572	333	11007	210239
Enthalpy	-9	0	0	-32	-4	-1	0	-1	-4	-23	-4	-10	0	0	-68	-33	0	-11	-68	-68	-11	-33	-77	-68	0	-8
Mass Flow																										
GLUCOSE	0	0	0	321	2303	493	0	0	21	128	17	43	0	0	0	0	0	43	0	0	43	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOSE	0	0	0	1880	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIGNIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2311	58	0	0	2311	2311	0	58	2311	2311	0	0
MICROO	0	0	0	0	0	0	148	690	0	0	0	0	0	0	690	690	0	690	690	690	690	690	690	690	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABIN	0	0	0	153	0	0	0	0	0	0	0	0	0	0	153	0	0	0	153	153	0	0	153	153	0	0
GALACT	0	0	0	65	0	0	0	0	0	0	0	0	0	0	65	0	0	0	65	65	0	0	65	65	0	0
MANNOS	0	0	0	16	0	0	0	0	0	0	0	0	0	0	16	0	0	0	16	16	0	0	16	16	0	0
ARABINA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANNAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALACTA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLOB	0	0	0	26	0	0	0	0	0	0	0	0	0	0	26	0	0	0	26	26	0	0	26	26	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
EXTRACT	0	0	0	767	0	0	0	0	5	30	1	3	0	0	800	33	0	3	800	800	3	33	800	800	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	403	0	0	7618	5	1	0	0	938	5713	961	2507	0	0	16859	8288	0	2507	16859	16859	2507	8268	17242	16859	20	382
N2	11495	123	573	0	0	0	0	0	0	0	0	0	2340	10922	0	573	0	0	0	0	0	0	10923	0	573	10922
CO2	3463	0	0	0	0	0	0	0	0	0	0	0	0	0	3	166	0	0	3	3	0	2	3303	3	164	3299
O2	722	35	162	0	0	0	0	0	0	0	0	0	660	3081	0	41	0	0	0	0	0	0	680	0	41	680
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	3	0	0	249	0	0	0	0	0	0	0	0	0	0	246	0	0	0	246	246	0	0	249	246	0	3
NH4SO4	0	0	0	0	0	0	0	0	26	158	8	20	0	0	178	178	0	20	178	178	20	178	178	178	0	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	16	0	0	24	0	0	0	0	0	0	0	0	0	0	7	0	0	0	7	7	0	0	24	7	0	16
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	0	0	0	44	0	0	0	0	0	0	0	0	0	0	44	0	0	0	44	44	0	0	44	44	0	0
CELLULA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFU-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	12	73	4	9	0	0	0	82	82	9	0	0	9	82	82	82	0	0
MGSO4-	0	0	0	0	0	0	0	0	3	18	1	2	0	0	0	20	20	2	0	0	2	20	20	20	0	0
HYDRAZI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-02	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DIAMM-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ORTHO-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

AREA 300 SHF

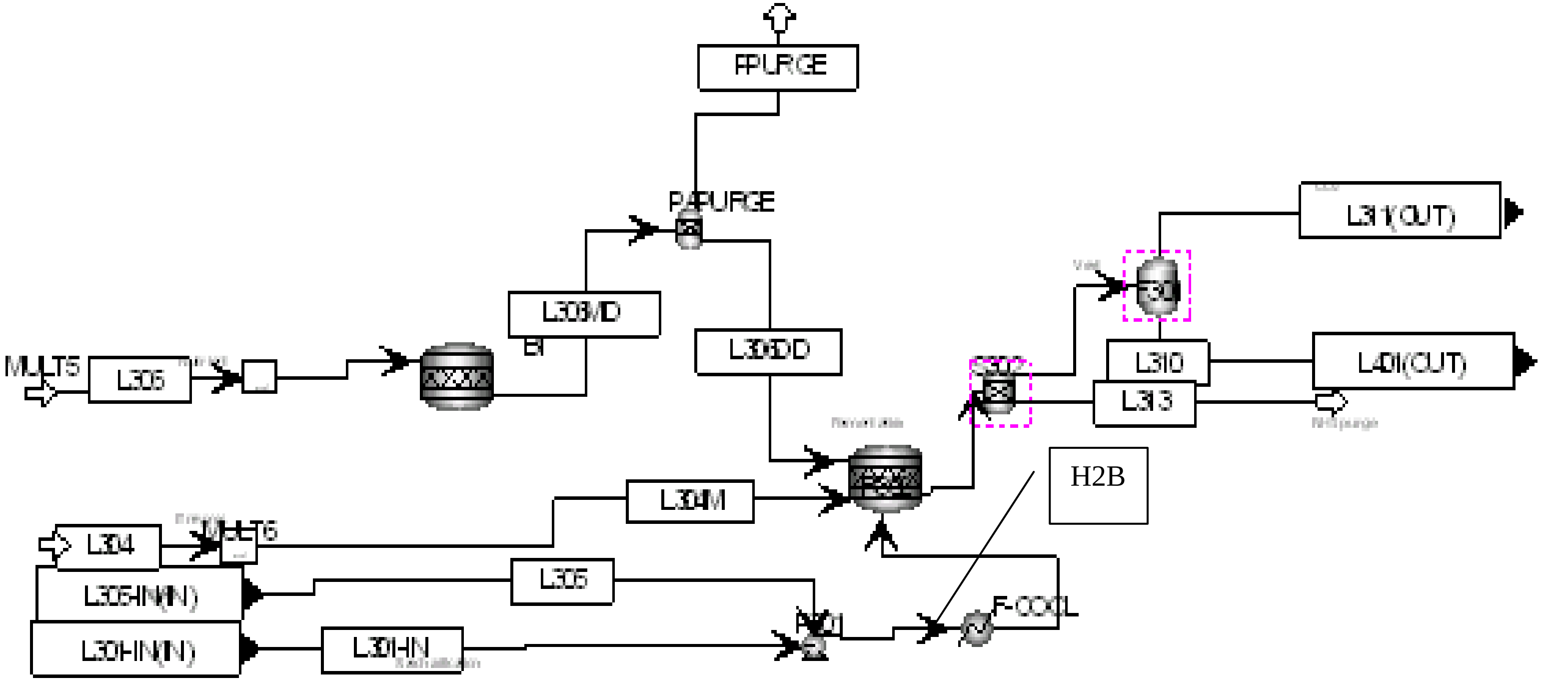

CO2
L311
Vent
F301
L312
L310
NH3 purge
Nutrient
L306
MULT9
L306A
Fermentation
R302
Seed
Enzymes
L305
L314B
L304
MULT10
L304A
H2A
COOLER
L301-IN
R301
Saccharification
L302
S301
L303
P301
L314
L307

AREA 300 SHF

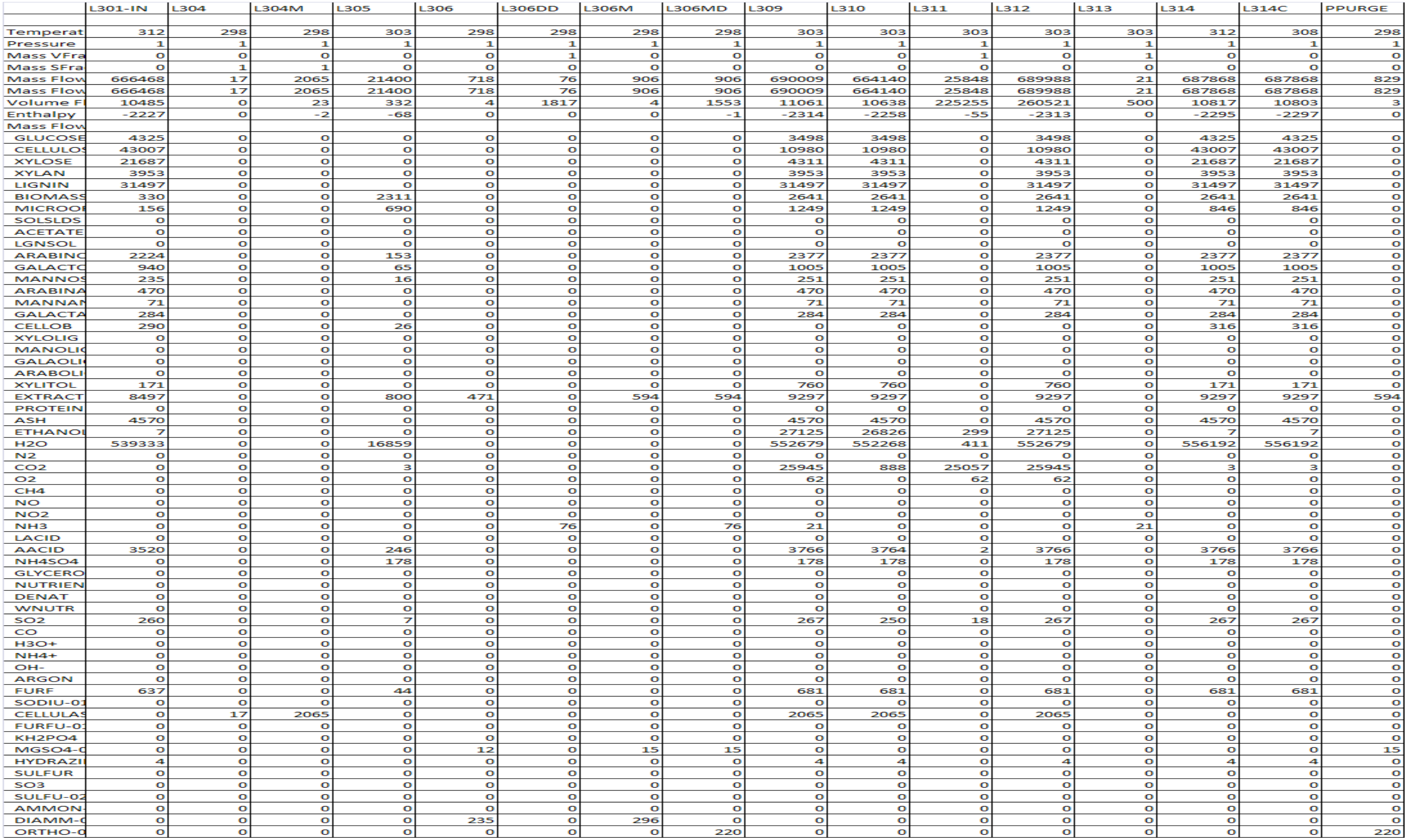


	L301-IN	L302	L303	L304	L304A	L305	L306	L306A	L307	L309	L310	L311	L312	L313	L314	L314B
Temperat	338	338	338	298	298	303	298	298	338	303	303	303	303	303	338	304
Pressure	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
Mass VFra	0	0	0	0	0	0	0	0	0	0	0	1	0	1	0	0
Mass SFra	0	0	0	1	1	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	632378	633804	570423	12	1426	60081	15	1786	63380	632290	609421	22760	632180	109	570423	630504
Mass Flow	632378	633804	570423	12	1426	60081	15	1786	63380	632290	609421	22760	632180	109	570423	630504
Volume F	10001	10092	9083	0	16	940	88	10764	1009	10027	9658	198241	229584	2639	9083	9884
Enthalpy	-2072	-2085	-1877	0	-1	-206	0	0	-209	-2106	-2056	-49	-2105	0	-1877	-2099
Mass Flow																
GLUCOSE	4172	44149	39734	0	0	0	0	0	4415	1391	1391	0	1391	0	39734	39734
CELLULOS	42989	7308	6577	0	0	731	0	0	731	7308	7308	0	7308	0	6577	7308
XYLOSE	25510	25510	22959	0	0	2551	0	0	2551	12946	12946	0	12946	0	22959	25510
XYLAN	3953	3953	3557	0	0	395	0	0	395	3953	3953	0	3953	0	3557	3953
LIGNIN	31497	31497	28348	0	0	3150	0	0	3150	31497	31497	0	31497	0	28348	31497
BIOMASS	49	49	44	0	0	348	0	0	5	392	392	0	392	0	44	392
MICROO	498	498	448	0	0	50	0	0	50	3983	3983	0	3983	0	448	498
UNKNOW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABIN	2377	2377	2140	0	0	238	0	0	238	2377	2377	0	2377	0	2140	2377
GALACT	1005	1005	905	0	0	101	0	0	101	1005	1005	0	1005	0	905	1005
MANNOS	251	251	226	0	0	25	0	0	25	251	251	0	251	0	226	251
ARABINA	470	470	423	0	0	47	0	0	47	470	470	0	470	0	423	470
MANNAN	71	71	64	0	0	7	0	0	7	71	71	0	71	0	64	71
GALACTA	284	284	255	0	0	28	0	0	28	284	284	0	284	0	255	284
CELLOB	316	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	75	75	68	0	0	8	0	0	8	334	334	0	334	0	68	75
EXTRACT	9271	9271	8344	0	0	927	11	1318	927	10589	10589	0	10589	0	8344	9271
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	4570	4570	4113	0	0	457	0	0	457	4570	4570	0	4570	0	4113	4570
ETHANO	6	6	5	0	0	1	0	0	1	23629	23376	253	23629	0	5	6
H2O	500278	496297	446667	0	0	50261	0	0	49630	498045	497684	361	498045	0	446667	496928
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	19	0	0	0	22898	800	22098	22898	0	0	19
O2	0	0	0	0	0	1	0	0	0	28	0	28	28	0	0	1
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	124	4	467	0	109	0	0	0	109	0	124
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	3766	3766	3390	0	0	376	0	0	377	3766	3764	2	3766	0	3390	3766
NH4SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	283	283	255	0	0	23	0	0	28	278	261	18	278	0	255	278
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	681	681	613	0	0	68	0	0	68	681	681	0	681	0	613	681
SO4--	1	1	1	0	0	0	0	0	0	1	1	0	1	0	1	1
SULFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CA(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULA	0	1426	1283	12	1426	143	0	0	143	1426	1426	0	1426	0	1283	1426
FURFU-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	1	0	0	0	1	1	0	1	0	0	1
MGSO4-	0	0	0	0	0	1	0	0	0	1	1	0	1	0	0	1
HYDRAZI	4	4	4	0	0	0	0	0	0	4	4	0	4	0	4	4
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

AREA 300 SSF

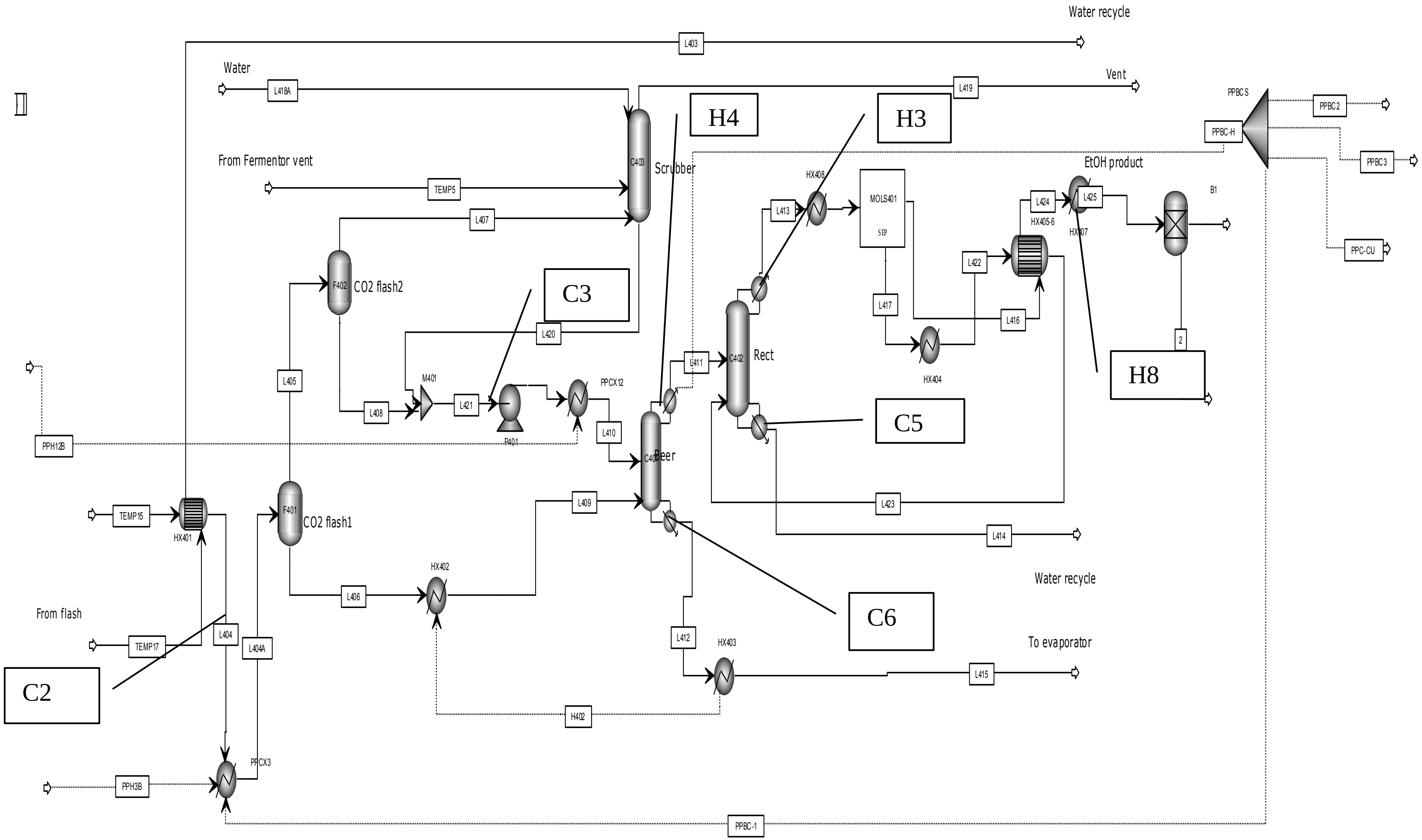

PPURGE
PAPURGE
L306MD
BI
L306DD
MULT5
L306
L304M
L304
MULT6
L305
L305HN(IN)
L301HN(IN)
L301HN
F-COOL
L311(OUT)
L310
L313
L401(OUT)
H2B

AREA 300 SSF



	L301-IN	L304	L304M	L305	L306	L306DD	L306M	L306MD	L309	L310	L311	L312	L313	L314	L314C	PPURGE
Temperat	312	298	298	303	298	298	298	298	303	303	303	303	303	312	308	298
Pressure	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
Mass VFra	0	0	0	0	0	1	0	0	0	0	1	0	1	0	0	0
Mass SFra	0	1	1	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	666468	17	2065	21400	718	76	906	906	690009	664140	25848	689988	21	687868	687868	829
Mass Flow	666468	17	2065	21400	718	76	906	906	690009	664140	25848	689988	21	687868	687868	829
Volume Fl	10485	0	23	332	4	1817	4	1553	11061	10638	225255	260521	500	10817	10803	3
Enthalpy	-2227	0	-2	-68	0	0	0	-1	-2314	-2258	-55	-2313	0	-2295	-2297	0
Mass Flow																
GLUCOSE	4325	0	0	0	0	0	0	0	3498	3498	0	3498	0	4325	4325	0
CELLULOS	43007	0	0	0	0	0	0	0	10980	10980	0	10980	0	43007	43007	0
XYLOSE	21687	0	0	0	0	0	0	0	4311	4311	0	4311	0	21687	21687	0
XYLAN	3953	0	0	0	0	0	0	0	3953	3953	0	3953	0	3953	3953	0
LIGNIN	31497	0	0	0	0	0	0	0	31497	31497	0	31497	0	31497	31497	0
BIOMASS	330	0	0	2311	0	0	0	0	2641	2641	0	2641	0	2641	2641	0
MICROOR	156	0	0	690	0	0	0	0	1249	1249	0	1249	0	846	846	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINO	2224	0	0	153	0	0	0	0	2377	2377	0	2377	0	2377	2377	0
GALACTO	940	0	0	65	0	0	0	0	1005	1005	0	1005	0	1005	1005	0
MANNOS	235	0	0	16	0	0	0	0	251	251	0	251	0	251	251	0
ARABINA	470	0	0	0	0	0	0	0	470	470	0	470	0	470	470	0
MANNAN	71	0	0	0	0	0	0	0	71	71	0	71	0	71	71	0
GALACTA	284	0	0	0	0	0	0	0	284	284	0	284	0	284	284	0
CELLOB	290	0	0	26	0	0	0	0	0	0	0	0	0	316	316	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLIC	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	171	0	0	0	0	0	0	0	760	760	0	760	0	171	171	0
EXTRACT	8497	0	0	800	471	0	594	594	9297	9297	0	9297	0	9297	9297	594
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	4570	0	0	0	0	0	0	0	4570	4570	0	4570	0	4570	4570	0
ETHANOL	7	0	0	0	0	0	0	0	27125	26826	299	27125	0	7	7	0
H2O	539333	0	0	16859	0	0	0	0	552679	552268	411	552679	0	556192	556192	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	3	0	0	0	0	25945	888	25057	25945	0	3	3	0
O2	0	0	0	0	0	0	0	0	62	0	62	62	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	76	0	76	21	0	0	0	21	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	3520	0	0	246	0	0	0	0	3766	3764	2	3766	0	3766	3766	0
NH4SO4	0	0	0	178	0	0	0	0	178	178	0	178	0	178	178	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	260	0	0	7	0	0	0	0	267	250	18	267	0	267	267	0
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	637	0	0	44	0	0	0	0	681	681	0	681	0	681	681	0
SODIU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULAS	0	17	2065	0	0	0	0	0	2065	2065	0	2065	0	0	0	0
FURFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MGSO4-0	0	0	0	0	12	0	15	15	0	0	0	0	0	0	0	15
HYDRAZI	4	0	0	0	0	0	0	0	4	4	0	4	0	4	4	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-02	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DIAMM-0	0	0	0	0	235	0	296	0	0	0	0	0	0	0	0	0
ORTHO-0	0	0	0	0	0	0	0	220	0	0	0	0	0	0	0	220

AREA 400


Water recycle
L403
Water
L418A
L419
Vent
PPBCS
PPBC2
PPBC-H
PPBC3
PPC-CU
H4
H3
From Fermentor vent
TEMP5
C403
Scrubber
HX408
MOLS401
SEP
EtOH product
L424
L425
B1
HX405-6
HX407
L413
L407
L422
F402
CO2 flash2
C3
L417
L416
L420
HX404
2
H8
L405
M401
PPCX12
L411
C402
Rect
L408
L421
P401
L410
C5
PPH12B
Beer
L423
L409
TEMP15
HX401
F401
CO2 flash1
L414
HX402
L406
Water recycle
C6
From flash
L404
TEMP17
L404A
L412
HX403
To evaporator
L415
C2
H402
PPCX3
PPH3B
PPBC-1

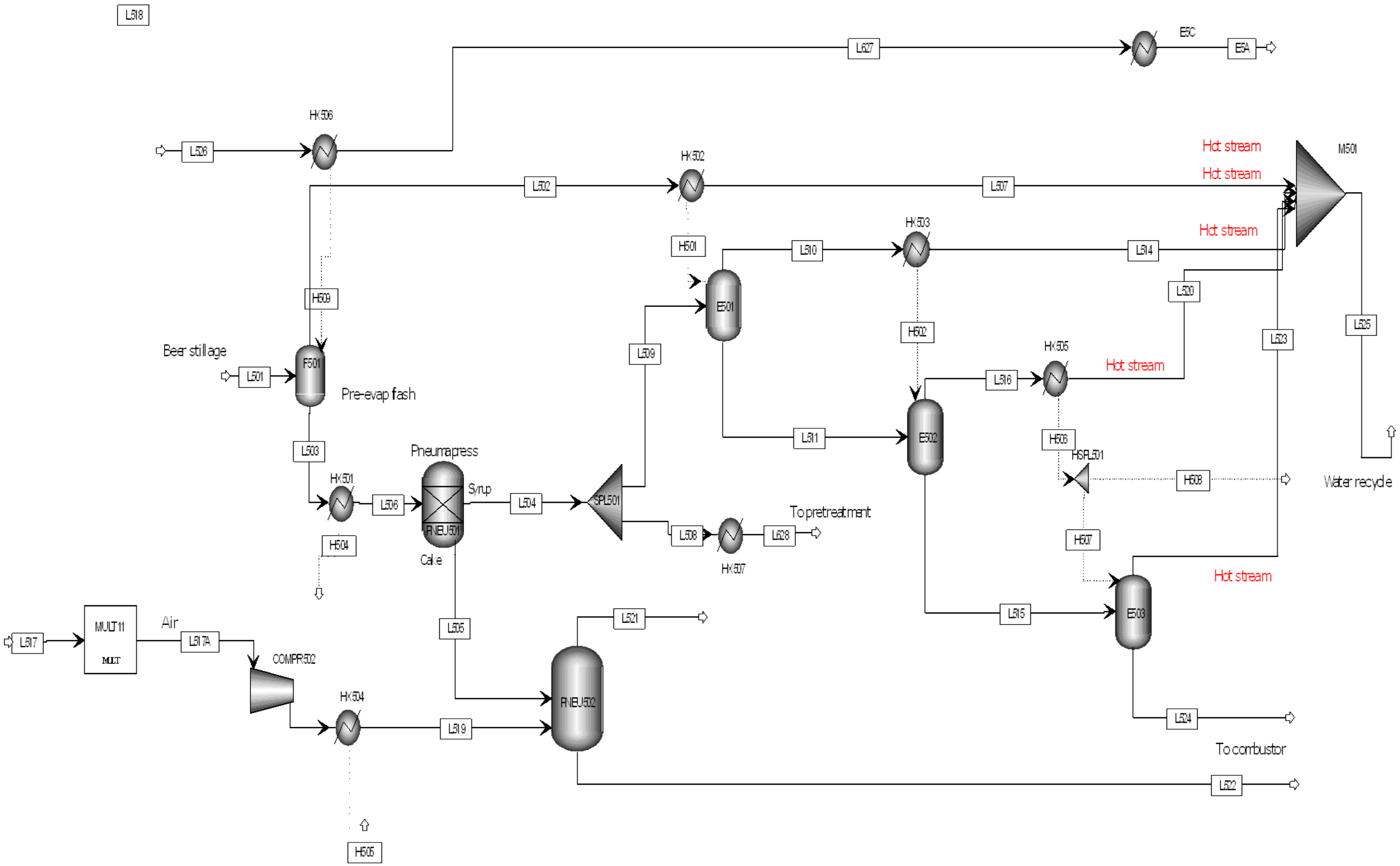
AREA 400 SHF

	1	2	L403	L404	L404A	L405	L406	L407	L408	L409	L410	L410B	L411	L412	L413	L414	L415	L416	L417	L418A	L419	L420	L421	L422	L423	L424	L425	L426	TEMP5	TEMP15	TEMP17
Temperat	298	298	372	319	363	359	359	311	311	381	358	295	375	388	359	388	366	389	389	286	287	295	295	308	353	358	298	393	303	303	374
Pressure	1	1	1	2	2	1	1	1	1	3	2	2	2	2	1	2	3	1	0	2	1	1	1	2	2	1	1	1	1	1	1
Mass VFra	0	1	1	0	0	1	0	1	0	0	0	0	1	0	1	0	0	1	1	0	1	0	0	0	0	1	0	1	1	0	1
Mass SFra	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	23666	272	33553	609421	609421	2786	606635	837	1949	606635	36678	36678	42832	600480	32400	18893	600480	23939	8461	34109	22977	34729	36678	8461	8461	23939	23939	32400	22760	609421	33553
Mass Flow	23666	272	33553	609421	609421	2786	606635	837	1949	606635	36678	36678	42832	600480	32400	18893	600480	23939	8461	34109	22977	34729	36678	8461	8461	23939	23939	32400	22760	609421	33553
Volume F	503	1355	452511	9721	14234	62857	10242	7554	35	10493	1062	617	524588	10371	284967	349	10123	212735	484718	569	201647	581	617	170	181	193339	506	307096	198241	9658	904043
Enthalpy	-34	0	-110	-2047	-2024	-7	-2017	-2	-6	-2005	-135	-137	-88	-2032	-44	-70	-2045	-28	-15	-130	-49	-131	-137	-18	-18	-29	-35	-44	-49	-2056	-101
Mass Flow																															
GLUCOSE	0	0	0	1391	1391	0	1391	0	0	1391	0	0	0	1391	0	0	1391	0	0	0	0	0	0	0	0	0	0	0	0	1391	0
CELLULOS	0	0	0	7308	7308	0	7308	0	0	7308	0	0	0	7308	0	0	7308	0	0	0	0	0	0	0	0	0	0	0	0	7308	0
XYLOSE	0	0	0	12946	12946	0	12946	0	0	12946	0	0	0	12946	0	0	12946	0	0	0	0	0	0	0	0	0	0	0	0	12946	0
XYLAN	0	0	0	3953	3953	0	3953	0	0	3953	0	0	0	3953	0	0	3953	0	0	0	0	0	0	0	0	0	0	0	0	3953	0
LIGNIN	0	0	0	31497	31497	0	31497	0	0	31497	0	0	0	31497	0	0	31497	0	0	0	0	0	0	0	0	0	0	0	0	31497	0
BIOMASS	0	0	0	392	392	0	392	0	0	392	0	0	0	392	0	0	392	0	0	0	0	0	0	0	0	0	0	0	0	392	0
MICROO	0	0	0	3983	3983	0	3983	0	0	3983	0	0	0	3983	0	0	3983	0	0	0	0	0	0	0	0	0	0	0	0	3983	0
UNKNOW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABIN	0	0	0	2377	2377	0	2377	0	0	2377	0	0	0	2377	0	0	2377	0	0	0	0	0	0	0	0	0	0	0	0	2377	0
GALACT	0	0	0	1005	1005	0	1005	0	0	1005	0	0	0	1005	0	0	1005	0	0	0	0	0	0	0	0	0	0	0	0	1005	0
MANNOS	0	0	0	251	251	0	251	0	0	251	0	0	0	251	0	0	251	0	0	0	0	0	0	0	0	0	0	0	0	251	0
ARABINA	0	0	0	470	470	0	470	0	0	470	0	0	0	470	0	0	470	0	0	0	0	0	0	0	0	0	0	0	0	470	0
MANNAN	0	0	0	71	71	0	71	0	0	71	0	0	0	71	0	0	71	0	0	0	0	0	0	0	0	0	0	0	0	71	0
GALACTA	0	0	0	284	284	0	284	0	0	284	0	0	0	284	0	0	284	0	0	0	0	0	0	0	0	0	0	0	0	284	0
CELLOB	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	334	334	0	334	0	0	334	0	0	0	334	0	0	334	0	0	0	0	0	0	0	0	0	0	0	0	334	0
EXTRACT	0	0	0	10589	10589	0	10589	0	0	10589	0	0	0	10589	0	0	10589	0	0	0	0	0	0	0	0	0	0	0	0	10589	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	4570	4570	0	4570	0	0	4570	0	0	0	4570	0	0	4570	0	0	0	0	0	0	0	0	0	0	0	0	4570	0
ETHANOL	23548	0	0	23376	23376	666	22710	55	611	22710	919	919	23558	71	29435	9	71	23548	5887	0	0	307	919	5887	5887	23548	23548	29435	253	23376	0
H2O	13	0	30925	497684	497684	1335	496349	19	1316	496349	35656	35656	18836	513169	2587	18823	513169	13	2574	34109	149	34340	35656	2574	2574	13	13	2587	361	497684	30925
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	104	0	0	800	800	759	42	754	5	42	62	62	104	0	104	0	0	104	0	0	22794	57	62	0	0	104	104	104	22098	800	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	28	0	0	0	0	0	0	0	28	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	0	0	565	3764	3764	5	3759	0	5	3759	7	7	59	3707	0	59	3707	0	0	0	0	2	7	0	0	0	0	0	2	3764	565
NH4SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	1	272	2037	261	261	21	240	9	11	240	34	34	274	0	274	0	0	274	0	0	5	22	34	0	0	274	274	274	18	261	2037
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	0	0	25	681	681	0	681	0	0	681	0	0	2	679	0	2	679	0	0	0	0	0	0	0	0	0	0	0	0	681	25
SO4--	0	0	0	1	1	0	1	0	0	1	0	0	0	1	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	1	0
SULFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CA(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULAS	0	0	0	1426	1426	0	1426	0	0	1426	0	0	0	1426	0	0	1426	0	0	0	0	0	0	0	0	0	0	0	0	1426	0
FURFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	1	1	0	1	0	0	1	0	0	0	1	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	1	0
MGSO4-	0	0	0	1	1	0	1	0	0	1	0	0	0	1	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	1	0
HYDRAZI	0	0	1	4	4	0	4	0	0	4	0	0	0	4	0	0	4	0	0	0	0	0	0	0	0	0	0	0	0	4	1
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

AREA 400 SSF

	1	2	L403	L404	L404B	L405	L406	L407	L408	L409	L410	L410B	L411	L412	L413	L414	L415	L416	L417	L418A	L419	L420	L421	L422	L423	L424	L425	L426	TEMP5	TEMP15	TEMP17
Temperat	298	298	372	319	362	359	359	311	311	381	345	295	375	388	359	388	366	389	389	286	287	295	295	308	353	358	298	393	303	303	374
Pressure	1	1	1	2	2	1	1	1	1	3	2	2	2	2	1	2	3	1	0	2	1	1	1	2	2	1	1	1	1	1	1
Mass VFra	0	1	0	0	0	1	0	1	0	0	0	0	1	0	1	0	0	1	1	0	1	0	0	0	0	1	0	1	1	0	1
Mass SFra	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	27164	261	33674	664140	664140	3212	660929	930	2282	660929	41743	41743	49169	653503	37193	21744	653503	27425	9768	38750	26066	39461	41743	9768	9768	27425	27425	37193	25848	664140	33674
Mass Flow	27164	261	33674	664140	664140	3212	660929	930	2282	660929	41743	41743	49169	653503	37193	21744	653503	27425	9768	38750	26066	39461	41743	9768	9768	27425	27425	37193	25848	664140	33674
Volume Fl	577	1300	409281	10707	15684	72546	11279	8393	41	11555	1009	702	602990	11404	327943	402	11131	243844	561897	646	228863	660	702	196	208	221501	581	353380	225255	10638	907468
Enthalpy	-39	0	-111	-2249	-2223	-8	-2215	-2	-7	-2201	-154	-156	-102	-2231	-51	-80	-2245	-33	-18	-147	-56	-149	-156	-21	-21	-33	-40	-50	-55	-2258	-102
Mass Flow																															
GLUCOSE	0	0	0	3498	3498	0	3498	0	0	3498	0	0	0	3498	0	0	3498	0	0	0	0	0	0	0	0	0	0	0	0	3498	0
CELLULOS	0	0	0	10980	10980	0	10980	0	0	10980	0	0	0	10980	0	0	10980	0	0	0	0	0	0	0	0	0	0	0	0	10980	0
XYLOSE	0	0	0	4311	4311	0	4311	0	0	4311	0	0	0	4311	0	0	4311	0	0	0	0	0	0	0	0	0	0	0	0	4311	0
XYLAN	0	0	0	3953	3953	0	3953	0	0	3953	0	0	0	3953	0	0	3953	0	0	0	0	0	0	0	0	0	0	0	0	3953	0
LIGNIN	0	0	0	31497	31497	0	31497	0	0	31497	0	0	0	31497	0	0	31497	0	0	0	0	0	0	0	0	0	0	0	0	31497	0
BIOMASS	0	0	0	2641	2641	0	2641	0	0	2641	0	0	0	2641	0	0	2641	0	0	0	0	0	0	0	0	0	0	0	0	2641	0
MICROOR	0	0	0	1249	1249	0	1249	0	0	1249	0	0	0	1249	0	0	1249	0	0	0	0	0	0	0	0	0	0	0	0	1249	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINO	0	0	0	2377	2377	0	2377	0	0	2377	0	0	0	2377	0	0	2377	0	0	0	0	0	0	0	0	0	0	0	0	2377	0
GALACTO	0	0	0	1005	1005	0	1005	0	0	1005	0	0	0	1005	0	0	1005	0	0	0	0	0	0	0	0	0	0	0	0	1005	0
MANNOS	0	0	0	251	251	0	251	0	0	251	0	0	0	251	0	0	251	0	0	0	0	0	0	0	0	0	0	0	0	251	0
ARABINA	0	0	0	470	470	0	470	0	0	470	0	0	0	470	0	0	470	0	0	0	0	0	0	0	0	0	0	0	0	470	0
MANNAN	0	0	0	71	71	0	71	0	0	71	0	0	0	71	0	0	71	0	0	0	0	0	0	0	0	0	0	0	0	71	0
GALACTA	0	0	0	284	284	0	284	0	0	284	0	0	0	284	0	0	284	0	0	0	0	0	0	0	0	0	0	0	0	284	0
CELLOB	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	760	760	0	760	0	0	760	0	0	0	760	0	0	760	0	0	0	0	0	0	0	0	0	0	0	0	760	0
EXTRACT	0	0	0	9297	9297	0	9297	0	0	9297	0	0	0	9297	0	0	9297	0	0	0	0	0	0	0	0	0	0	0	0	9297	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	4570	4570	0	4570	0	0	4570	0	0	0	4570	0	0	4570	0	0	0	0	0	0	0	0	0	0	0	0	4570	0
ETHANOL	27032	0	0	26826	26826	798	26028	62	737	26028	1097	1097	27043	81	33790	11	81	27032	6758	0	0	360	1097	6758	6758	27032	27032	33790	299	26826	0
H2O	15	0	31045	552268	552268	1543	550725	21	1522	550725	40535	40535	21684	569575	3025	21669	569575	15	3010	38750	169	39012	40535	3010	3010	15	15	3025	411	552268	31045
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	115	0	0	888	888	844	44	838	6	44	71	71	115	0	115	0	0	115	0	0	25830	65	71	0	0	115	115	115	25057	888	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	62	0	0	0	0	0	0	0	62	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	0	0	565	3764	3764	6	3759	0	6	3759	7	7	61	3705	0	61	3705	0	0	0	0	2	7	0	0	0	0	0	2	3764	565
NH4SO4	0	0	0	178	178	0	178	0	0	178	0	0	0	178	0	0	178	0	0	0	0	0	0	0	0	0	0	0	0	178	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	1	261	2038	250	250	21	229	9	12	229	34	34	263	0	263	0	0	263	0	0	5	22	34	0	0	263	263	263	18	250	2038
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	0	0	25	681	681	0	681	0	0	681	0	0	3	679	0	3	679	0	0	0	0	0	0	0	0	0	0	0	0	681	25
SULFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CA(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULAS	0	0	0	2065	2065	0	2065	0	0	2065	0	0	0	2065	0	0	2065	0	0	0	0	0	0	0	0	0	0	0	0	2065	0
FURFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MGSO4-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HYDRAZI	0	0	1	4	4	0	4	0	0	4	0	0	0	4	0	0	4	0	0	0	0	0	0	0	0	0	0	0	0	4	1
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-02	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DIAMM-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ORTHO-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

AREA 500


L518
L627
E5C
E5A
HX506
L526
HX502
L502
L507
Hot stream
Hot stream
M501
H501
L510
HX503
L514
Hot stream
L520
H509
E501
L509
H502
L523
L525
Beer stillage
F501
L501
Pre-evap flash
HX505
L516
Hot stream
L503
HX501
Pneumapress
L511
E502
H506
HSPL501
H508
Water recycle
Syrup
L506
L504
SPL501
PNEU501
To pretreatment
L508
L528
H504
HX507
H507
Cake
Hot stream
E503
L515
Air
L521
MULT11
MULT
L517
L517A
L505
COMPR502
HX504
PNEU502
L519
L524
To combustor
L522
H505

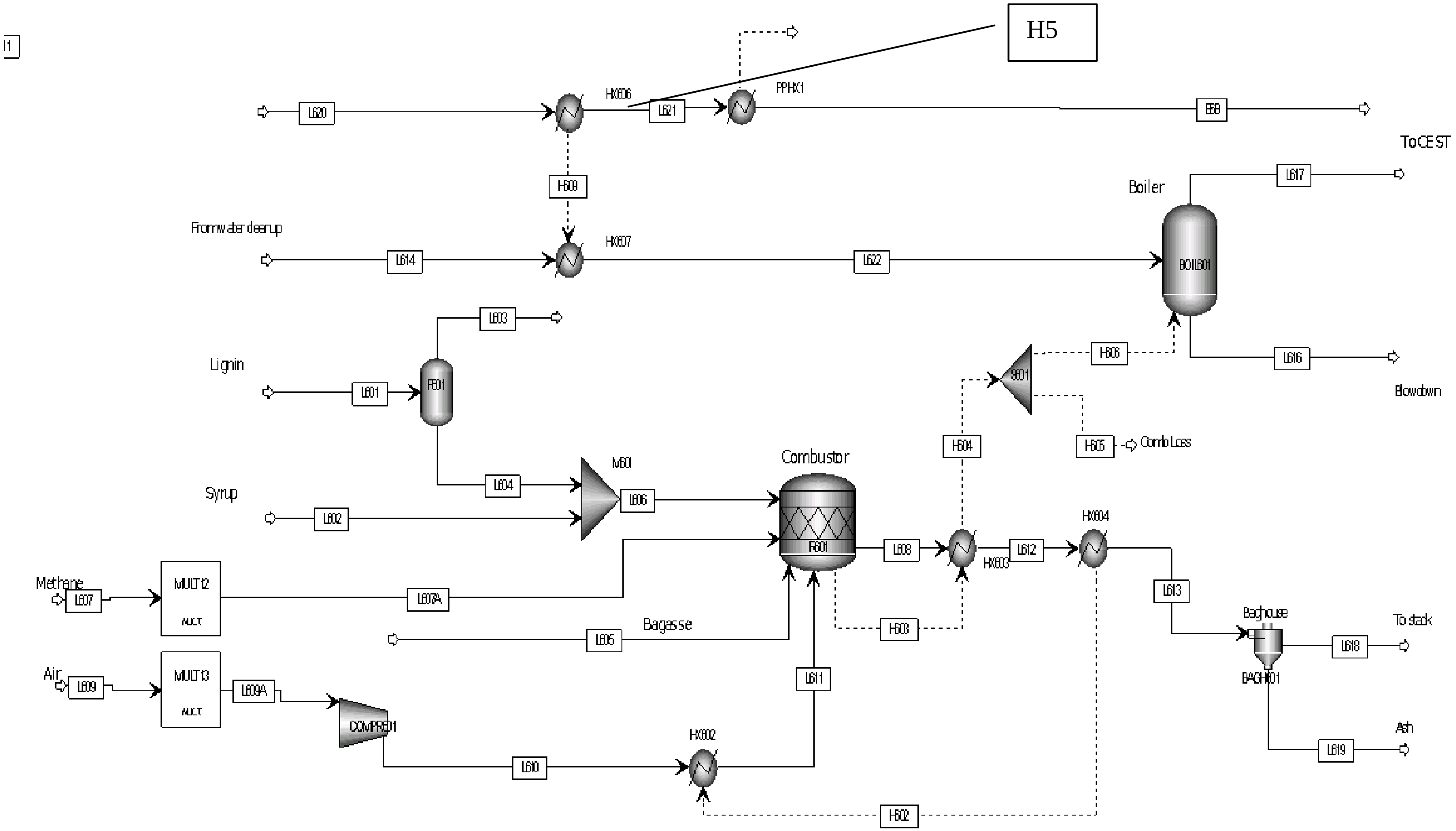
AREA 500 SHF

	E5A	L501	L502	L503	L504	L505	L506	L507	L508	L509	L510	L511	L514	L515	L516	L517	L517A	L518	L519	L520	L521	L522	L523	L524	L525	L526	L627	L628
Temperat	373	366	349	349	343	343	343	349	343	343	339	339	339	335	335	298	298	647	313	335	338	338	329	329	373	598	373	298
Pressure	1	3	0	0	3	3	0	0	3	3	0	0	0	0	0	1	1	10	10	0	3	3	0	0	1	4	1	1
Mass VFra	0	0	1	0	0	0	0	0	0	0	1	0	0	0	1	1	1	1	1	0	1	0	1	0	0	1	0	0
Mass SFra	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	40080	600480	66052	534429	426254	108174	534429	66052	106564	319691	67646	252045	67646	183370	68675	78	9537	9537	9537	68675	10015	107696	70522	112848	272895	40080	40080	106564
Mass Flow	40080	600480	66052	534429	426254	108174	534429	66052	106564	319691	67646	252045	67646	183370	68675	78	9537	9537	9537	68675	10015	107696	70522	112848	272895	40080	40080	106564
Volume Fl	697	10123	4508750	8480	7166	1293	8454	1130	1791	5374	6746190	4213	1184	3087	8341350	1096	134728	30848	14902	1165	51409	1282	10936800	1839	1569540	408175	697	1756
Enthalpy	-148	-2045	-210	-1809	-1546	-265	-1811	-246	-387	-1160	-215	-908	-253	-652	-219	0	0	1	0	-257	-1	-264	-225	-389	-981	-123	-148	-391
Mass Flow																												
GLUCOSE	0	1391	0	1391	1252	139	1391	0	313	939	0	939	0	939	0	0	0	0	0	0	0	139	0	939	0	0	0	313
CELLULOS	0	7308	0	7308	146	7162	7308	0	37	110	0	110	0	110	0	0	0	0	0	0	0	7162	0	110	0	0	0	37
XYLOSE	0	12946	0	12946	11652	1295	12946	0	2913	8739	0	8739	0	8739	0	0	0	0	0	0	0	1295	0	8739	0	0	0	2913
XYLAN	0	3953	0	3953	79	3874	3953	0	20	59	0	59	0	59	0	0	0	0	0	0	0	3874	0	59	0	0	0	20
LIGNIN	0	31497	0	31497	630	30867	31497	0	157	472	0	472	0	472	0	0	0	0	0	0	0	30867	0	472	0	0	0	157
BIOMASS	0	392	0	392	196	196	392	0	49	147	0	147	0	147	0	0	0	0	0	0	0	196	0	147	0	0	0	49
MICROOR	0	3983	0	3983	1991	1991	3983	0	498	1494	0	1494	0	1494	0	0	0	0	0	0	0	1991	0	1494	0	0	0	498
UNKNOW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINO	0	2377	0	2377	2140	238	2377	0	535	1605	0	1605	0	1605	0	0	0	0	0	0	0	238	0	1605	0	0	0	535
GALACTO	0	1005	0	1005	905	101	1005	0	226	678	0	678	0	678	0	0	0	0	0	0	0	101	0	678	0	0	0	226
MANNOS	0	251	0	251	226	25	251	0	57	170	0	170	0	170	0	0	0	0	0	0	0	25	0	170	0	0	0	57
ARABINA	0	470	0	470	9	461	470	0	2	7	0	7	0	7	0	0	0	0	0	0	0	461	0	7	0	0	0	2
MANNAN	0	71	0	71	1	69	71	0	0	1	0	1	0	1	0	0	0	0	0	0	0	69	0	1	0	0	0	0
GALACTA	0	284	0	284	6	278	284	0	1	4	0	4	0	4	0	0	0	0	0	0	0	278	0	4	0	0	0	1
CELLOB	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	334	0	334	300	33	334	0	75	225	0	225	0	225	0	0	0	0	0	0	0	33	0	225	0	0	0	75
EXTRACT	0	10589	0	10589	212	10377	10589	0	53	159	0	159	0	159	0	0	0	0	0	0	0	10377	0	159	0	0	0	53
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	4570	0	4570	91	4479	4570	0	23	69	0	69	0	69	0	0	0	0	0	0	0	4479	0	69	0	0	0	23
ETHANOL	2	71	46	25	22	2	25	46	6	17	13	3	13	1	3	0	0	0	0	3	0	2	0	0	63	2	2	6
H2O	39801	513169	65708	447461	402715	44746	447461	65708	100679	302036	67292	234744	67292	166470	68274	0	0	0	0	68274	477	44269	70015	96455	271289	39801	39801	100679
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	60	7316	7316	7316	0	7314	1	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	18	2221	2221	2221	0	2221	1	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	121	3707	285	3422	3080	342	3422	285	770	2310	326	1984	326	1606	379	0	0	0	0	379	2	340	475	1130	1465	121	121	770
NH4SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	132	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	132	132	0
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	6	679	12	667	600	67	667	12	150	450	15	436	15	416	20	0	0	0	0	20	0	67	31	385	77	6	6	150
SO4--	0	1	0	1	1	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CA(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULAS	0	1426	0	1426	0	1426	1426	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1426	0	0	0	0	0	0
FURFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	1	0	1	0	1	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0
MGSO4-0	0	1	0	1	0	1	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0
HYDRAZI	18	4	0	4	0	4	4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4	0	0	0	18	18	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

AREA 500 SSF

	L501	L502	L503	L504	L505	L506	L507	L508	L509	L510	L511	L514	L515	L516	L517	L517A	L518	L519	L520	L521	L522	L523	L524	L525	L526	L627	L627C	L628
Temperat	366	349	349	343	343	343	349	343	343	339	339	339	335	335	298	298	647	313	335	338	338	329	329	373	571	373	373	298
Pressure	3	0	0	3	3	0	0	3	3	0	0	0	0	0	1	1	10	10	0	3	3	0	0	1	5	1	1	1
Mass VFra	0	1	0	0	0	0	0	0	0	1	0	0	0	1	1	1	1	1	0	1	0	1	0	0	1	0	0	0
Mass SFra	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	653503	69396	584107	468330	115776	584107	69396	117083	351248	71241	280007	71241	207578	72430	78	9537	9537	9537	72430	10026	115287	74595	132982	287662	42063	42063	42063	117083
Mass Flow	653503	69396	584107	468330	115776	584107	69396	117083	351248	71241	280007	71241	207578	72430	78	9537	9537	9537	72430	10026	115287	74595	132982	287662	42063	42063	42063	117083
Volume Fl	11131	4737590	9374	7908	1437	9345	1187	1977	5931	7105950	4707	1247	3539	8799240	1096	134728	30848	14902	1229	51558	1426	11570900	2198	1662170	401819	731	731	1938
Enthalpy	-2245	-221	-1998	-1715	-286	-2001	-259	-429	-1286	-227	-1021	-266	-751	-231	0	0	1	0	-271	-1	-284	-238	-473	-1034	-129	-156	-156	-434
Mass Flow																												
GLUCOSE	3498	0	3498	3148	350	3498	0	787	2361	0	2361	0	2361	0	0	0	0	0	0	0	350	0	2361	0	0	0	0	787
CELLULO	10980	0	10980	220	10760	10980	0	55	165	0	165	0	165	0	0	0	0	0	0	0	10760	0	165	0	0	0	0	55
XYLOSE	4311	0	4311	3879	431	4311	0	970	2910	0	2910	0	2910	0	0	0	0	0	0	0	431	0	2910	0	0	0	0	970
XYLAN	3953	0	3953	79	3874	3953	0	20	59	0	59	0	59	0	0	0	0	0	0	0	3874	0	59	0	0	0	0	20
LIGNIN	31497	0	31497	630	30867	31497	0	157	472	0	472	0	472	0	0	0	0	0	0	0	30867	0	472	0	0	0	0	157
BIOMASS	2641	0	2641	1321	1321	2641	0	330	991	0	991	0	991	0	0	0	0	0	0	0	1321	0	991	0	0	0	0	330
MICROO	1249	0	1249	625	625	1249	0	156	468	0	468	0	468	0	0	0	0	0	0	0	625	0	468	0	0	0	0	156
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINO	2377	0	2377	2140	238	2377	0	535	1605	0	1605	0	1605	0	0	0	0	0	0	0	238	0	1605	0	0	0	0	535
GALACTO	1005	0	1005	905	101	1005	0	226	678	0	678	0	678	0	0	0	0	0	0	0	101	0	678	0	0	0	0	226
MANNOS	251	0	251	226	25	251	0	57	170	0	170	0	170	0	0	0	0	0	0	0	25	0	170	0	0	0	0	57
ARABINA	470	0	470	9	461	470	0	2	7	0	7	0	7	0	0	0	0	0	0	0	461	0	7	0	0	0	0	2
MANNAN	71	0	71	1	69	71	0	0	1	0	1	0	1	0	0	0	0	0	0	0	69	0	1	0	0	0	0	0
GALACTA	284	0	284	6	278	284	0	1	4	0	4	0	4	0	0	0	0	0	0	0	278	0	4	0	0	0	0	1
CELLOB	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	760	0	760	684	76	760	0	171	513	0	513	0	513	0	0	0	0	0	0	0	76	0	513	0	0	0	0	171
EXTRACT	9297	0	9297	186	9111	9297	0	46	139	0	139	0	139	0	0	0	0	0	0	0	9111	0	139	0	0	0	0	46
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	4570	0	4570	91	4479	4570	0	23	69	0	69	0	69	0	0	0	0	0	0	0	4479	0	69	0	0	0	0	23
ETHANOL	81	52	29	26	3	29	52	7	20	16	4	16	1	4	0	0	0	0	4	0	3	1	0	72	2	2	2	7
H2O	569575	69064	500512	450461	50051	500512	69064	112615	337845	70905	266940	70905	194885	72055	0	0	0	0	72055	489	49562	74132	120753	286156	41781	41781	41781	112615
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	60	7316	7316	7316	0	7314	2	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	18	2221	2221	2221	0	2220	1	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	3705	269	3436	3092	344	3436	269	773	2319	306	2013	306	1660	353	0	0	0	0	353	2	342	436	1224	1364	119	119	119	773
NH4SO4	178	0	178	0	178	178	0	0	0	0	0	0	0	0	0	0	0	0	0	0	178	0	0	0	0	0	0	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	138	138	138	0
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	679	11	668	601	67	668	11	150	451	14	437	14	419	18	0	0	0	0	18	0	67	27	392	70	6	6	6	150
SODIU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CA(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULAS	2065	0	2065	0	2065	2065	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2065	0	0	0	0	0	0	0
FURFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MGSO4-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HYDRAZI	4	0	4	0	4	4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4	0	0	0	18	18	18	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-02	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DIAMM-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ORTHO-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

AREA 600



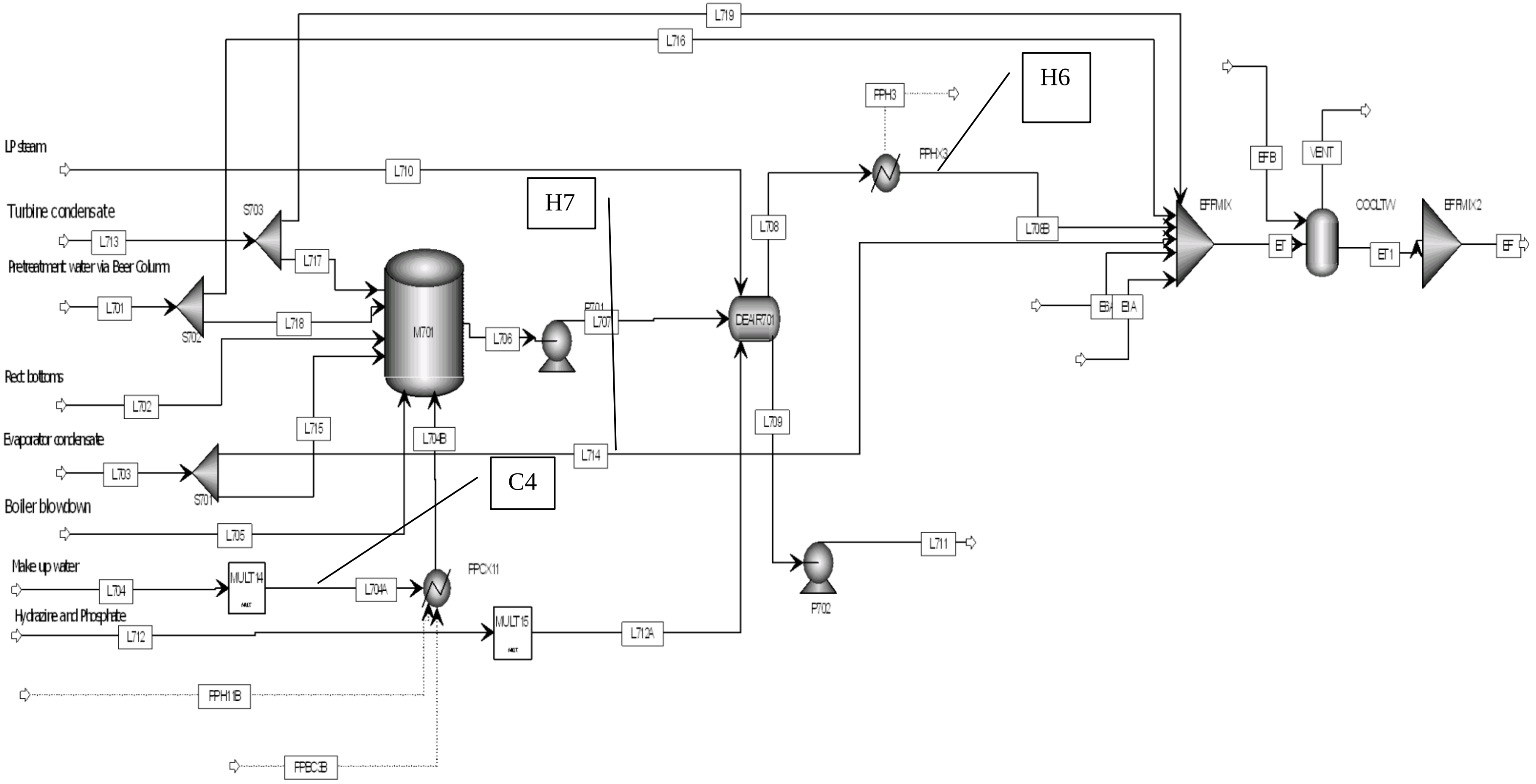
AREA 600 SHF

	E6B	L601	L602	L603	L604	L605	L606	L607	L607A	L608	L609	L609A	L610	L611	L612	L613	L614	L616	L617	L618	L619	L620	L621	L622
Temperat	375	338	329	338	338		332	298	298	1143	298	298	300	506	552	430	412	1029	1029	430	430	733	465	467
Pressure	13	3	0	1	1		1	1	1	1	1	1	1	1	1	1	96	87	87	1	1	13	13	96
Mass VFra	0	0	0	1	0		0	1	1	1	1	1	1	1	1	1	0	0	1	1	0	1	0	0
Mass SFra	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0
Mass Flow	25607	107696	112848	2	107694	0	220543	4	430	774030	4500	553057	553057	553057	774030	774030	277217	0	277217	765648	8382	25607	25607	277217
Mass Flow	25607	107696	112848	2	107694		220543	4	430	774030	4500	553057	553057	553057	774030	774030	277217	0	277217	765648	8382	25607	25607	277217
Volume Fl	446	1282	1839	33	1282		3112	89	10913	46917600	64071	7874400	7806650	13192200	22575800	17594200	4956	0	242931	17594100	65	108968	488	5263
Enthalpy	-95	-264	-389	0	-264	0	-652	0	0	-707	0	-23	-23	5	-859	-887	-1015	0	-789	-870	-17	-77	-92	-999
Mass Flow																								
GLUCOSE	0	139	939	0	139		1078	0	0	11	0	0	0	0	11	11	0	0	0	11	0	0	0	0
CELLULOS	0	7162	110	0	7162		7272	0	0	73	0	0	0	0	73	73	0	0	0	1	72	0	0	0
XYLOSE	0	1295	8739	0	1295		10033	0	0	100	0	0	0	0	100	100	0	0	0	100	0	0	0	0
XYLAN	0	3874	59	0	3874		3933	0	0	39	0	0	0	0	39	39	0	0	0	0	39	0	0	0
LIGNIN	0	30867	472	0	30867		31340	0	0	313	0	0	0	0	313	313	0	0	0	4	310	0	0	0
BIOMASS	0	196	147	0	196		343	0	0	3	0	0	0	0	3	3	0	0	0	0	3	0	0	0
MICROO	0	1991	1494	0	1991		3485	0	0	3485	0	0	0	0	3485	3485	0	0	0	42	3443	0	0	0
UNKNOW	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINO	0	238	1605	0	238		1842	0	0	18	0	0	0	0	18	18	0	0	0	18	0	0	0	0
GALACTO	0	101	678	0	101		779	0	0	8	0	0	0	0	8	8	0	0	0	8	0	0	0	0
MANNOS	0	25	170	0	25		195	0	0	2	0	0	0	0	2	2	0	0	0	2	0	0	0	0
ARABINA	0	461	7	0	461		468	0	0	5	0	0	0	0	5	5	0	0	0	0	5	0	0	0
MANNAN	0	69	1	0	69		71	0	0	1	0	0	0	0	1	1	0	0	0	0	1	0	0	0
GALACTA	0	278	4	0	278		282	0	0	3	0	0	0	0	3	3	0	0	0	0	3	0	0	0
CELLOB	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLI	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	33	225	0	33		259	0	0	3	0	0	0	0	3	3	0	0	0	3	0	0	0	0
EXTRACT	0	10377	159	0	10377		10536	0	0	10536	0	0	0	0	10536	10536	0	0	0	10536	0	0	0	0
PROTEIN	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	4479	69	0	4479		4547	0	0	4547	0	0	0	0	4547	4547	0	0	0	55	4493	0	0	0
ETHANOL	1	2	0	0	2		2	0	0	0	0	0	0	0	0	0	13	0	13	0	0	1	1	13
H2O	25428	44269	96455	0	44268		140724	0	0	191103	59	7190	7190	7190	191103	191103	275287	0	275288	191103	0	25428	25428	275288
N2	0	1	0	1	0		0	0	0	419848	3414	419549	419549	419549	419848	419848	0	0	0	419848	0	0	0	0
CO2	0	0	0	0	0		0	0	0	135962	0	0	0	0	135962	135962	0	0	0	135962	0	0	0	0
O2	0	1	0	1	0		0	0	0	7895	1028	126318	126318	126318	7895	7895	0	0	0	7895	0	0	0	0
CH4	0	0	0	0	0		0	4	430	4	0	0	0	0	4	4	0	0	0	4	0	0	0	0
NO	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	77	340	1130	0	340		1470	0	0	15	0	0	0	0	15	15	839	0	839	15	0	77	77	839
NH4SO4	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCERO	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	84	0	0	0	0		0	0	0	31	0	0	0	0	31	31	913	0	913	31	0	84	84	913
CO	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	4	67	385	0	67		451	0	0	5	0	0	0	0	5	5	41	0	41	5	0	4	4	41
SO4--	0	0	0	0	0		1	0	0	1	0	0	0	0	1	1	0	0	0	1	0	0	0	0
SULFU-01	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULAS	0	1426	0	0	1426		1426	0	0	14	0	0	0	0	14	14	0	0	0	0	14	0	0	0
FURFU-01	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	1	0	0	1		1	0	0	1	0	0	0	0	1	1	0	0	0	1	0	0	0	0
MGSO4-0	0	1	0	0	1		1	0	0	1	0	0	0	0	1	1	0	0	0	1	0	0	0	0
HYDRAZI	11	4	0	0	4		4	0	0	4	0	0	0	0	4	4	124	0	124	4	0	11	11	124
SULFUR	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

AREA 600 SSF

	L601	L602	L603	L604	L605	L606	L607	L607A	L608	L609	L609A	L610	L611	L612	L613	L614	L616	L617	L618	L619	L620	L621	L621C	L622
Mass Flow	60489	130747		60487		191233								9516	9594	290939	0		9594			26874	26874	290939
Volume Fl	880	2174		880		3048								23	21	5201	0		21			513	468	5513
Heat Cap	3	4		3		4								0	0	4	2793		0			4	4	4
Conductiv	0	0		0		0								0	0	1	1		0			1	1	1
Viscosity	0	0		0		0								718	0	0	1		0			0	0	0
Surface Te	65	67		65		66								55	74	51	0		74			39	58	39
MWMX	21	19		21		20								187	176	18	18		176			18	18	18
Temperat	338	329	338	338		331	298	298	1143	298	298	300	518	552	430	412	987	987	430	430	701	465	375	465
Pressure	3	0	1	1		1	1	1	1	1	1	1	1	1	1	96	87	87	1	1	13	13	13	96
Mass VFra	0	0	1	0		0	1	1	1	1	1	1	1	1	1	0	0	1	1	0	1	0	0	0
Mass SFra	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0
Mass Flow	115287	132982	2	115285	0	248267	7	860	802185	4500	553057	553057	553057	802185	802185	290939	0	290938	796105	6080	26874	26874	26874	290939
Mass Flow	115287	132982	2	115285		248267	7	860	802185	4500	553057	553057	553057	802185	802185	290939	0	290938	796105	6080	26874	26874	26874	290939
Volume Fl	1426	2198	37	1426		3618	178	21826	49675100	64071	7874400	7806650	13511300	23906900	18630400	5201	0	243257	18630400	40	107673	513	468	5513
Enthalpy	-284	-473	0	-284	0	-757	0	-1	-805	0	-23	-23	7	-966	-996	-1065	0	-836	-982	-14	-81	-97	-99	-1049
Mass Flow																								
GLUCOSE	350	2361	0	350		2711	0	0	27	0	0	0	0	27	27	0	0	0	27	0	0	0	0	0
CELLULOS	10760	165	0	10760		10925	0	0	109	0	0	0	0	109	109	0	0	0	1	108	0	0	0	0
XYLOSE	431	2910	0	431		3341	0	0	33	0	0	0	0	33	33	0	0	0	33	0	0	0	0	0
XYLAN	3874	59	0	3874		3933	0	0	39	0	0	0	0	39	39	0	0	0	0	39	0	0	0	0
LIGNIN	30867	472	0	30867		31340	0	0	313	0	0	0	0	313	313	0	0	0	4	310	0	0	0	0
BIOMASS	1321	991	0	1321		2311	0	0	23	0	0	0	0	23	23	0	0	0	0	23	0	0	0	0
MICROO	625	468	0	625		1093	0	0	1093	0	0	0	0	1093	1093	0	0	0	14	1079	0	0	0	0
SOLSLDS	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABIN	238	1605	0	238		1842	0	0	18	0	0	0	0	18	18	0	0	0	18	0	0	0	0	0
GALACT	101	678	0	101		779	0	0	8	0	0	0	0	8	8	0	0	0	8	0	0	0	0	0
MANNOS	25	170	0	25		195	0	0	2	0	0	0	0	2	2	0	0	0	2	0	0	0	0	0
ARABINA	461	7	0	461		468	0	0	5	0	0	0	0	5	5	0	0	0	0	5	0	0	0	0
MANNAN	69	1	0	69		71	0	0	1	0	0	0	0	1	1	0	0	0	0	1	0	0	0	0
GALACTA	278	4	0	278		282	0	0	3	0	0	0	0	3	3	0	0	0	0	3	0	0	0	0
CELLOB	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLI	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	76	513	0	76		589	0	0	6	0	0	0	0	6	6	0	0	0	6	0	0	0	0	0
EXTRACT	9111	139	0	9111		9251	0	0	9251	0	0	0	0	9251	9251	0	0	0	9251	0	0	0	0	0
PROTEIN	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	4479	69	0	4479		4547	0	0	4547	0	0	0	0	4547	4547	0	0	0	55	4493	0	0	0	0
ETHANO	3	0	0	3		3	0	0	0	0	0	0	0	0	0	16	0	16	0	0	1	1	1	16
H2O	49562	120753	0	49561		170315	0	0	222578	59	7190	7190	7190	222578	222578	288983	0	288983	222578	0	26693	26693	26693	288983
N2	2	0	1	0		0	0	0	420230	3414	419549	419549	419549	420230	420230	0	0	0	420230	0	0	0	0	0
CO2	0	0	0	0		0	0	0	141219	0	0	0	0	141219	141219	0	0	0	141219	0	0	0	0	0
O2	1	0	1	0		0	0	0	2386	1028	126318	126318	126318	2386	2386	0	0	0	2386	0	0	0	0	0
CH4	0	0	0	0		0	7	860	9	0	0	0	0	9	9	0	0	0	9	0	0	0	0	0
NO	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	342	1224	0	342		1566	0	0	16	0	0	0	0	16	16	824	0	824	16	0	76	76	76	824
NH4SO4	178	0	0	178		178	0	0	178	0	0	0	0	178	178	0	0	0	178	0	0	0	0	0
GLYCERO	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0		0	0	0	62	0	0	0	0	62	62	953	0	953	62	0	88	88	88	953
CO	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	67	392	0	67		459	0	0	5	0	0	0	0	5	5	39	0	39	5	0	4	4	4	39
CA(OH)2	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULA	2065	0	0	2065		2065	0	0	21	0	0	0	0	21	21	0	0	0	0	20	0	0	0	0
FURFU-0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MGSO4-	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HYDRAZ	4	0	0	4		4	0	0	4	0	0	0	0	4	4	124	0	124	4	0	11	11	11	124
SULFUR	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-02	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DIAMM-	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ORTHO-	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

AREA 700


L719
L716
FPH3
H6
FPHX3
LP steam
L710
H7
Turbine condensate
S703
L713
L717
L708
L708B
EFRMIX
EFB
VENT
COOLTW
EFRMIX2
ET
ET1
EF
Pretreatment water via Beer Column
L701
S702
L718
M701
L706
P701
L707
DEAIR701
EIA
Rect bottoms
L702
L715
L704B
L709
Evaporator condensate
L703
S701
L714
C4
Boiler blowdown
L705
L711
Make up water
L704
MULT14
L704A
FPCX11
P702
Hydrazine and Phosphate
L712
MULT15
L712A
FPH11B
FPEC3B

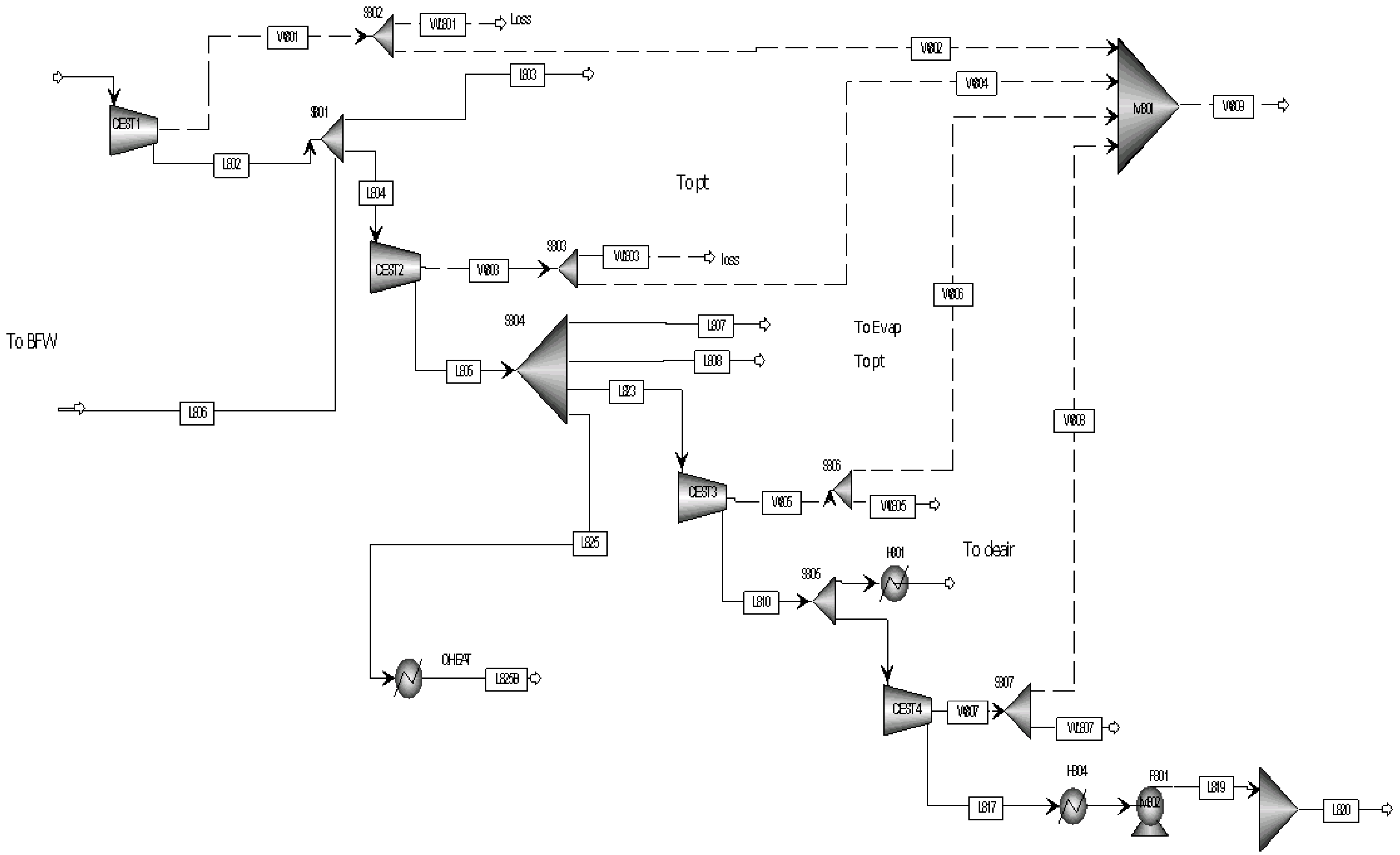
AREA 700 SHF

	E1A	E6A	EF	EFB	ET	ET1	L701	L702	L703	L704	L704A	L704B	L705	L706	L707	L708	L708B	L709	L710	L711	L712	L712A	L713	L714	L715	L716	L717	L718	L719	VENT
Temperat	373	375	308	373	373	308	372	388	373	298	298	364	1029	414	414	410	375	410	390	412	298	298	318	373	373	372	318	372	318	308
Pressure	1	13	0	1	1	0	1	2	1	1	1	1	87	2	3	3	1	3	2	96	1	1	4	1	1	1	4	1	4	0
Mass VFra	1	0	0	0	0	0	1	0	0	0	0	0	0	0	0	1	1	0	0	0	0	0	0	0	0	1	0	1	0	1
Mass SFra	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	9450	25607	297124	40080	286594	297124	33553	18893	272895	1500	184352	184352	0	287292	287292	24401	24401	277217	14203	277217	1	123	4734	218316	54579	5033	947	28520	3788	29549
Mass Flow	9450	25607	297124	40080	286594	297124	33553	18893	272895	1500	184352	184352	0	287292	287292	24401	24401	277217	14203	277217	1	123	4734	218316	54579	5033	947	28520	3788	29549
Volume Fl	146585	446	4991	697	2162060	4980	452611	349	1569540	25	3081	3186	0	5173	5174	219845	671058	4975	250	4956	0	2	80	1255630	313908	67892	16	384720	64	11824600
Enthalpy	-32	-95	-1120	-148	-1018	-1120	-110	-70	-981	-6	-699	-687	0	-1050	-1050	-75	-76	-1015	-52	-1015	0	0	-18	-785	-196	-16	-4	-93	-14	-92
Mass Flow																														
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIGNIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MICROO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UNKNOW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALACTO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANNOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANNAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALACTA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLOB	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	1	24	2	62	24	0	9	63	0	0	0	0	22	22	10	10	13	1	13	0	0	0	50	13	0	0	0	0	39
H2O	9433	25428	295413	39801	283769	295413	30925	18823	271289	1500	184352	184352	0	284659	284659	23476	23476	275287	14104	275287	0	0	4701	217031	54258	4639	940	26286	3761	28156
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	1	77	1426	121	1386	1426	565	59	1465	0	0	0	0	835	835	39	39	839	43	839	0	0	14	1172	293	85	3	481	11	82
NH4SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	15	84	150	132	1286	150	2037	0	0	0	0	0	0	1735	1735	869	869	913	47	913	0	0	16	0	0	306	3	1732	12	1269
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	0	4	76	6	70	76	25	2	77	0	0	0	0	39	39	1	1	41	2	41	0	0	1	62	15	4	0	21	1	1
SODIU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULAS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MGSO4-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HYDRAZI	0	11	36	18	20	36	1	0	0	0	0	0	0	1	1	7	7	124	6	124	1	123	2	0	0	0	0	1	2	2
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

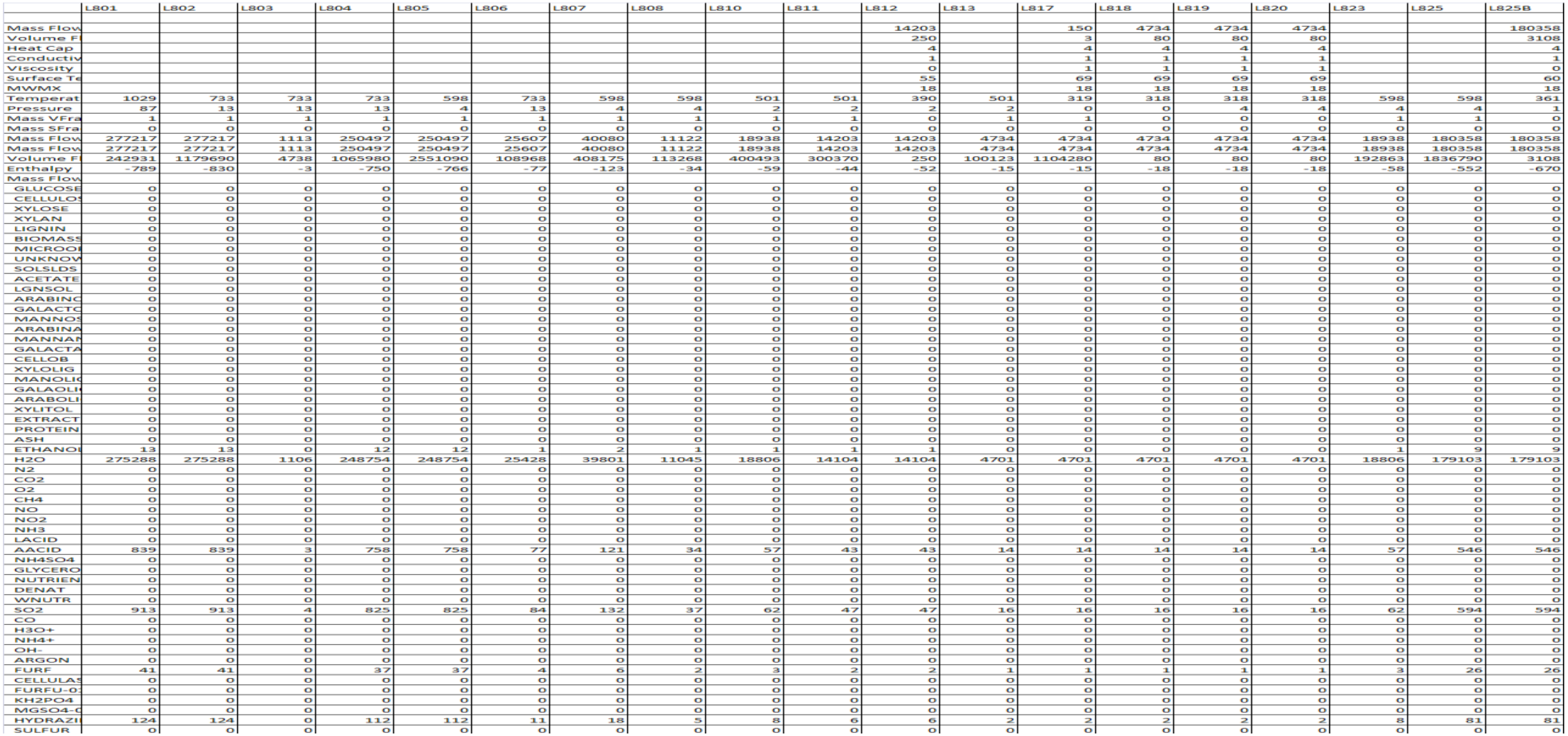
AREA 700 SSF

	E1	E2	EF	EFB	ET	ETC	L701	L702	L703	L704	L704A	L704B	L705	L706	L707	L708	L708B	L709	L710	L711	L712	L712A	L713	L714	L715	L716	L717	L718	L719	VENT
Mass Flow	3829	26874	310256	42063	220418	310256	17762	21744	228550	1550	190497	190497	0	299390	299390			290939	14906	290939	1	123	4969	182840	45710	2664	994	15098	3975	
Volume F	67	468	5201	731	3834	5201	308	402	3976	26	3184	3267	0	5341	5341			5222	262	5201	0	2	84	3181	795	46	17	262	67	
Heat Cap	4	4	4	4	4	4	4	5	4	4	4	4	2793	4	4			4	4	4	3	3	4	4	4	4	4	4	4	
Conductiv	1	1	1	1	1	1	0	1	1	1	1	1	1	1	1			1	1	1	1	1	1	1	1	0	1	0	1	
Viscosity	0	0	1	0	0	1	0	0	0	1	1	0	1	0	0			0	0	0	1	1	1	0	0	0	1	0	1	
Surface Te	58	58	71	58	58	71	58	55	58	73	73	62	0	52	52			51	55	51	67	67	69	58	58	58	69	58	69	
MWMX	18	18	18	18	18	18	18	18	18	18	18	18	18	18	18			18	18	18	32	32	18	18	18	18	18	18	18	
Temperat	373	375	308	373	373	308	372	388	373	298	298	353	987	404	404	410	375	410	389	412	298	298	318	373	373	372	318	372	318	308
Pressure	1	13	0	1	1	0	1	2	1	1	1	1	87	2	3	3	1	3	2	96	1	1	4	1	1	1	4	1	4	0
Mass VFra	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	1	0	0	0	0	0	0	0	0	0	0	0	0	1
Mass SFra	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	9489	26874	310256	42063	298998	310256	33674	21744	287662	1550	190497	190497	0	299390	299390	23480	23480	290939	14906	290939	1	123	4969	230129	57532	5051	994	28623	3975	30805
Mass Flow	9489	26874	310256	42063	298998	310256	33674	21744	287662	1550	190497	190497	0	299390	299390	23480	23480	290939	14906	290939	1	123	4969	230129	57532	5051	994	28623	3975	30805
Volume F	158902	468	5201	731	2185630	5201	409287	402	1662170	26	3184	3267	0	5341	5341	211586	645848	5222	262	5201	0	2	84	1329730	332433	61393	17	347894	67	12357900
Enthalpy	-32	-99	-1170	-156	-1063	-1170	-111	-80	-1034	-6	-722	-712	0	-1097	-1097	-72	-73	-1066	-55	-1065	0	0	-19	-827	-207	-17	-4	-95	-15	-96
Mass Flow																														
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIGNIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MICROO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALACTO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANNOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANNAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALACTA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLOB	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	1	28	2	70	28	0	11	72	0	0	0	0	26	26	11	11	16	1	16	0	0	0	57	14	0	0	0	0	45
H2O	9471	26693	308635	41781	296290	308635	31045	21669	286156	1550	190497	190497	0	296773	296773	22596	22596	288983	14806	288983	0	0	4935	228925	57231	4657	987	26388	3948	29436
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	1	76	1342	119	1300	1342	565	61	1364	0	0	0	0	817	817	35	35	824	42	824	0	0	14	1091	273	85	3	480	11	77
NH4SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	16	88	147	138	1254	147	2038	0	0	0	0	0	0	1735	1735	832	832	953	49	953	0	0	16	0	0	306	3	1732	13	1245
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	0	4	69	6	64	69	25	3	70	0	0	0	0	37	37	0	0	39	2	39	0	0	1	56	14	4	0	21	1	1
SULFU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CA(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULAS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFU-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MGSO4-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HYDRAZI	0	11	36	18	20	36	1	0	0	0	0	0	0	1	1	6	6	124	6	124	1	123	2	0	0	0	0	1	2	2
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-02	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DIAMM-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ORTHO-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

AREA 800

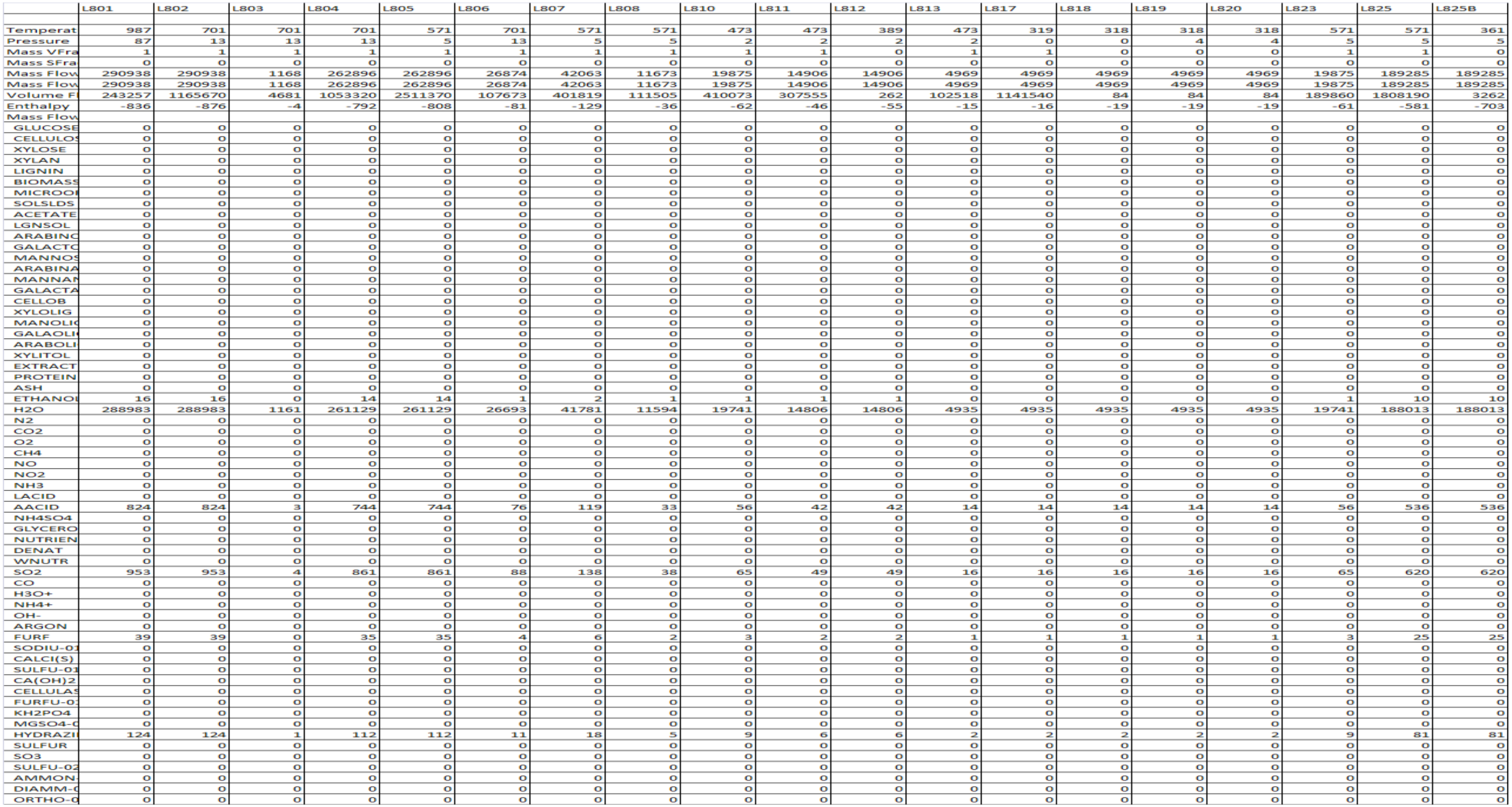

S802
V801
VL801
Loss
V802
CEST1
S801
L803
L802
V804
M801
V809
L804
To pt
CEST2
V803
S803
VL803
loss
V806
S804
L807
L808
To Evap
To pt
To BFW
L805
L823
L806
V808
S806
CEST3
V805
VL805
L825
To deair
H801
S805
L810
OHEAT
L825B
S807
CEST4
V807
VL807
H804
F801
M802
L819
L817
L820

AREA 800 SHF



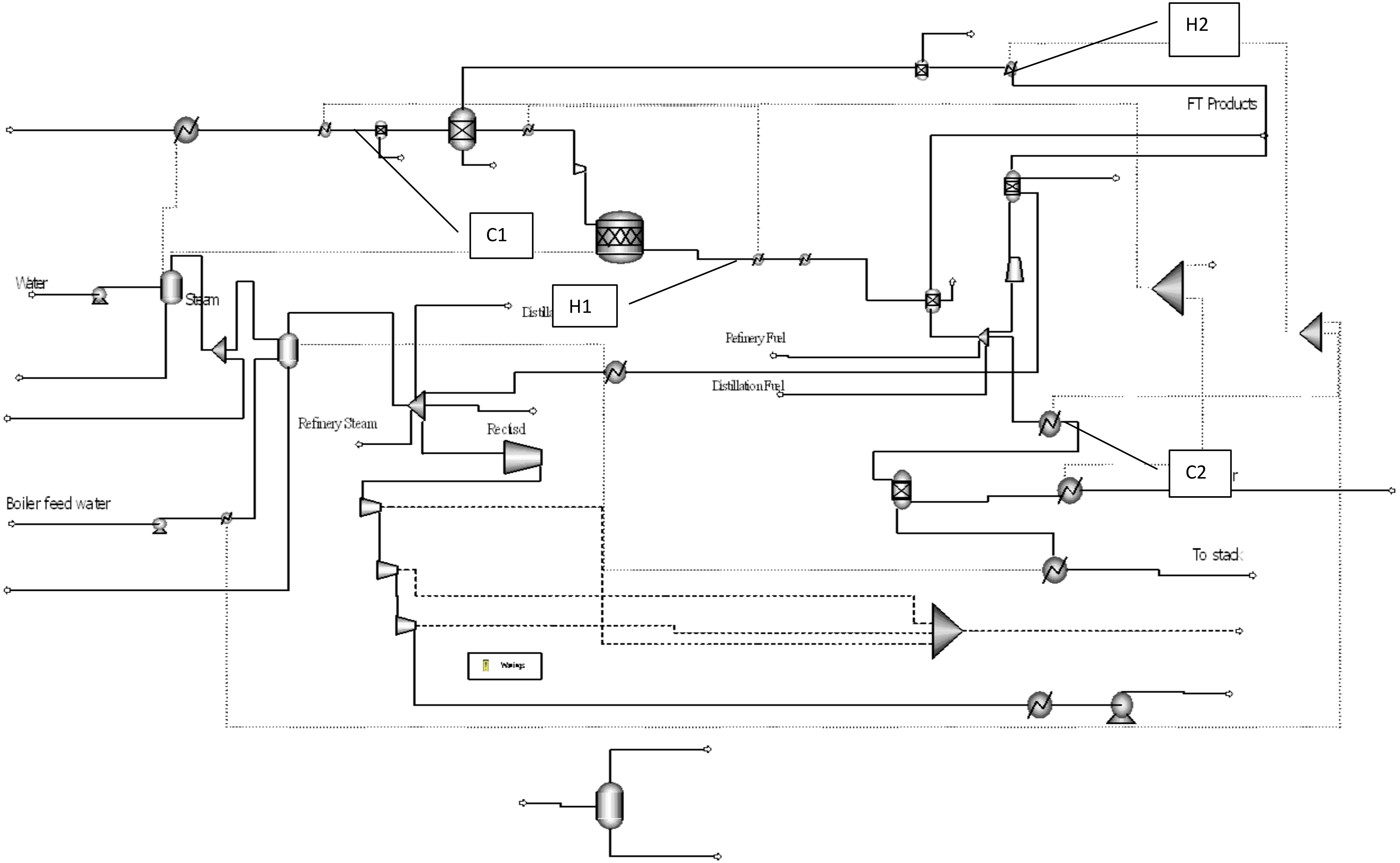
	L801	L802	L803	L804	L805	L806	L807	L808	L810	L811	L812	L813	L817	L818	L819	L820	L823	L825	L825B
Mass Flow											14203		150	4734	4734	4734			180358
Volume F											250		3	80	80	80			3108
Heat Cap											4		4	4	4	4			4
Conductiv											1		1	1	1	1			1
Viscosity											0		1	1	1	1			0
Surface Te											55		69	69	69	69			60
MWMX											18		18	18	18	18			18
Temperat	1029	733	733	733	598	733	598	598	501	501	390	501	319	318	318	318	598	598	361
Pressure	87	13	13	13	4	13	4	4	2	2	2	2	0	0	4	4	4	4	1
Mass VFra	1	1	1	1	1	1	1	1	1	1	0	1	1	0	0	0	1	1	0
Mass SFra	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	277217	277217	1113	250497	250497	25607	40080	11122	18938	14203	14203	4734	4734	4734	4734	4734	18938	180358	180358
Mass Flow	277217	277217	1113	250497	250497	25607	40080	11122	18938	14203	14203	4734	4734	4734	4734	4734	18938	180358	180358
Volume F	242931	1179690	4738	1065980	2551090	108968	408175	113268	400493	300370	250	100123	1104280	80	80	80	192863	1836790	3108
Enthalpy	-789	-830	-3	-750	-766	-77	-123	-34	-59	-44	-52	-15	-15	-18	-18	-18	-58	-552	-670
Mass Flow																			
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIGNIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MICROO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UNKNOW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALACTO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANNOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANNAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALACTA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLOB	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	13	13	0	12	12	1	2	1	1	1	1	0	0	0	0	0	1	9	9
H2O	275288	275288	1106	248754	248754	25428	39801	11045	18806	14104	14104	4701	4701	4701	4701	4701	18806	179103	179103
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	839	839	3	758	758	77	121	34	57	43	43	14	14	14	14	14	57	546	546
NH4SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	913	913	4	825	825	84	132	37	62	47	47	16	16	16	16	16	62	594	594
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	41	41	0	37	37	4	6	2	3	2	2	1	1	1	1	1	3	26	26
CELLULAS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFU-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MGSO4-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HYDRAZI	124	124	0	112	112	11	18	5	8	6	6	2	2	2	2	2	8	81	81
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

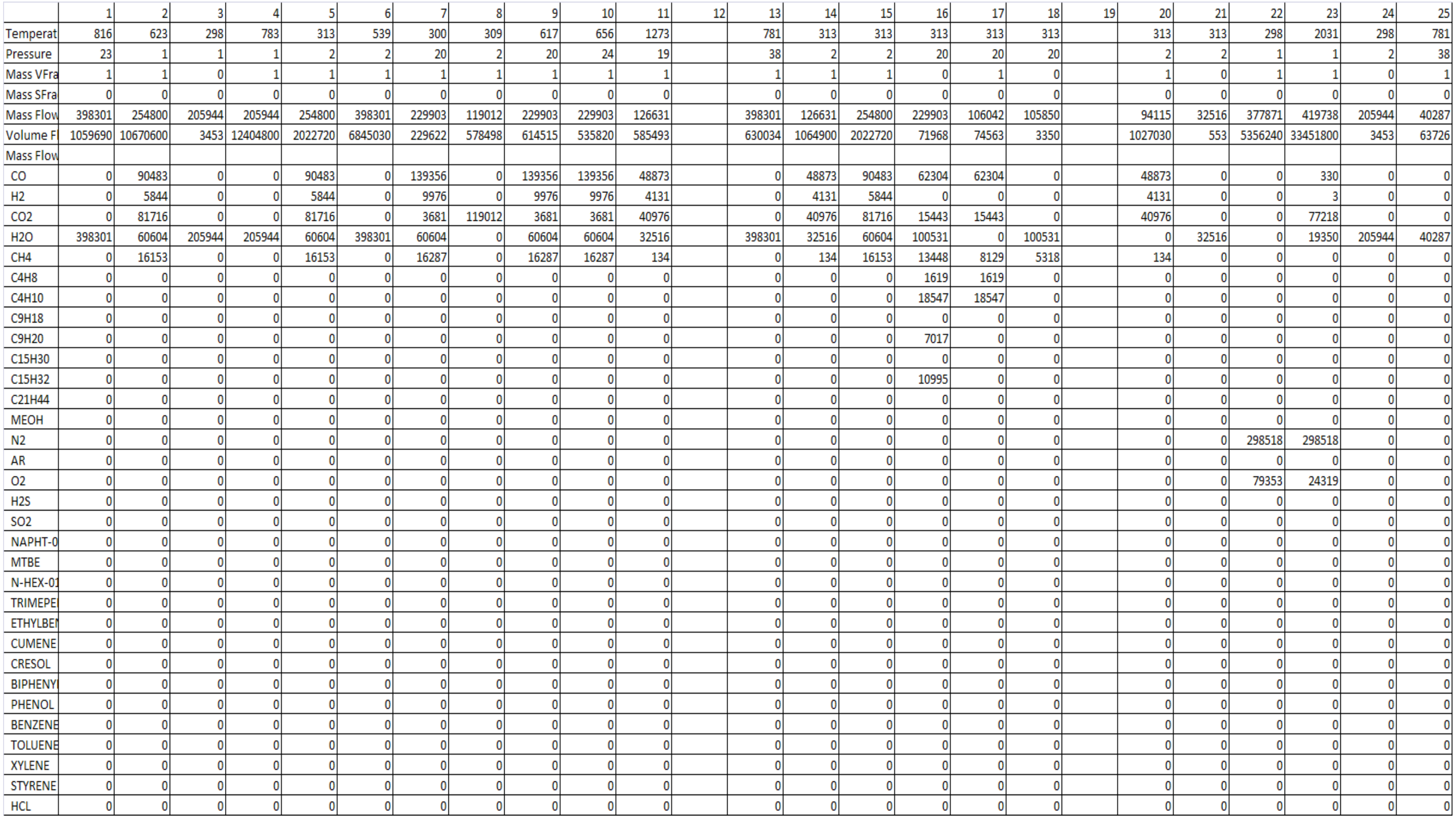
AREA 800 SSF



	L801	L802	L803	L804	L805	L806	L807	L808	L810	L811	L812	L813	L817	L818	L819	L820	L823	L825	L825B
Temperat	987	701	701	701	571	701	571	571	473	473	389	473	319	318	318	318	571	571	361
Pressure	87	13	13	13	5	13	5	5	2	2	2	2	0	0	4	4	5	5	5
Mass VFra	1	1	1	1	1	1	1	1	1	1	0	1	1	0	0	0	1	1	0
Mass SFra	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow	290938	290938	1168	262896	262896	26874	42063	11673	19875	14906	14906	4969	4969	4969	4969	4969	19875	189285	189285
Mass Flow	290938	290938	1168	262896	262896	26874	42063	11673	19875	14906	14906	4969	4969	4969	4969	4969	19875	189285	189285
Volume F	243257	1165670	4681	1053320	2511370	107673	401819	111505	410073	307555	262	102518	1141540	84	84	84	189860	1808190	3262
Enthalpy	-836	-876	-4	-792	-808	-81	-129	-36	-62	-46	-55	-15	-16	-19	-19	-19	-61	-581	-703
Mass Flow																			
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIGNIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MICROO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SOLSLDS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LGNSOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALACT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANNO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANNA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALACTA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLOB	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MANOLI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GALAOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANO	16	16	0	14	14	1	2	1	1	1	1	0	0	0	0	0	1	10	10
H2O	288983	288983	1161	261129	261129	26693	41781	11594	19741	14806	14806	4935	4935	4935	4935	4935	19741	188013	188013
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	824	824	3	744	744	76	119	33	56	42	42	14	14	14	14	14	56	536	536
NH4SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLYCERO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NUTRIEN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DENAT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
WNUTR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	953	953	4	861	861	88	138	38	65	49	49	16	16	16	16	16	65	620	620
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURF	39	39	0	35	35	4	6	2	3	2	2	1	1	1	1	1	3	25	25
SODIU-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI(S)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CA(OH)2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFU-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MGSO4-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HYDRAZ	124	124	1	112	112	11	18	5	9	6	6	2	2	2	2	2	9	81	81
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFU-02	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DIAMM-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ORTHO-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

APPENDIX 1.2 GFT PROCESS FLOW DIAGRAM





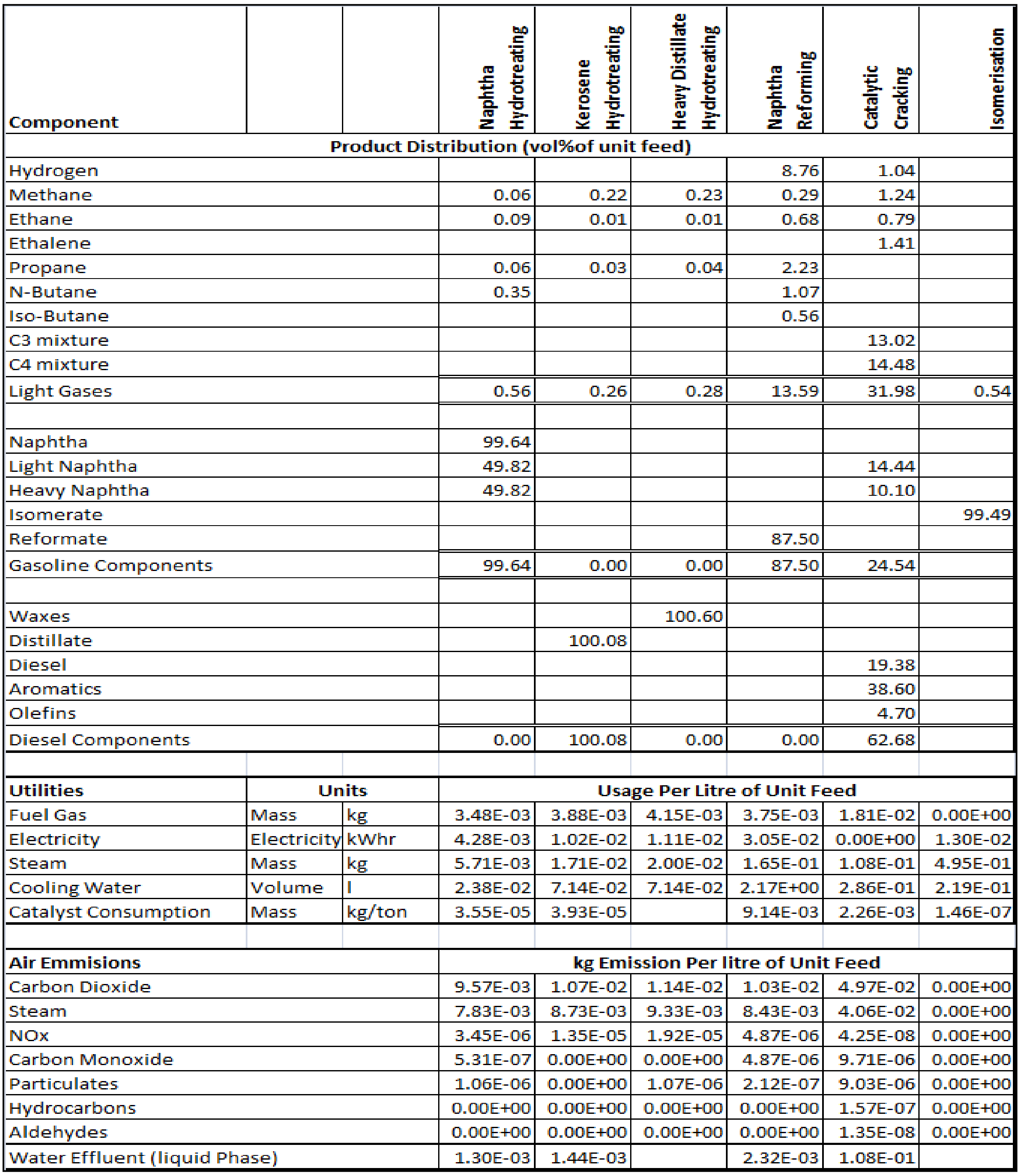
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25
Temperat	816	623	298	783	313	539	300	309	617	656	1273		781	313	313	313	313	313		313	313	298	2031	298	781
Pressure	23	1	1	1	2	2	20	2	20	24	19		38	2	2	20	20	20		2	2	1	1	2	38
Mass VFra	1	1	0	1	1	1	1	1	1	1	1		1	1	1	0	1	0		1	0	1	1	0	1
Mass SFra	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
Mass Flow	398301	254800	205944	205944	254800	398301	229903	119012	229903	229903	126631		398301	126631	254800	229903	106042	105850		94115	32516	377871	419738	205944	40287
Volume Fl	1059690	10670600	3453	12404800	2022720	6845030	229622	578498	614515	535820	585493		630034	1064900	2022720	71968	74563	3350		1027030	553	5356240	33451800	3453	63726
Mass Flow																									
CO	0	90483	0	0	90483	0	139356	0	139356	139356	48873		0	48873	90483	62304	62304	0		48873	0	0	330	0	0
H2	0	5844	0	0	5844	0	9976	0	9976	9976	4131		0	4131	5844	0	0	0		4131	0	0	3	0	0
CO2	0	81716	0	0	81716	0	3681	119012	3681	3681	40976		0	40976	81716	15443	15443	0		40976	0	0	77218	0	0
H2O	398301	60604	205944	205944	60604	398301	60604	0	60604	60604	32516		398301	32516	60604	100531	0	100531		0	32516	0	19350	205944	40287
CH4	0	16153	0	0	16153	0	16287	0	16287	16287	134		0	134	16153	13448	8129	5318		134	0	0	0	0	0
C4H8	0	0	0	0	0	0	0	0	0	0	0		0	0	0	1619	1619	0		0	0	0	0	0	0
C4H10	0	0	0	0	0	0	0	0	0	0	0		0	0	0	18547	18547	0		0	0	0	0	0	0
C9H18	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
C9H20	0	0	0	0	0	0	0	0	0	0	0		0	0	0	7017	0	0		0	0	0	0	0	0
C15H30	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
C15H32	0	0	0	0	0	0	0	0	0	0	0		0	0	0	10995	0	0		0	0	0	0	0	0
C21H44	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
MEOH	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	298518	298518	0	0
AR	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	79353	24319	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
NAPHT-0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
MTBE	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
N-HEX-01	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
TRIMEPE	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
ETHYLBE	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
CUMENE	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
CRESOL	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
BIPHENY	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
PHENOL	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
BENZENE	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
TOLUENE	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
XYLENE	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
STYRENE	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0		0	0	0	0	0	0

	26	27	28	29	30	31	32	34	35	37	38	39	40	41	44	45	46	47	48	49	52	AIR	BFW	BLOWDOV	FLUEGAS	FTPROD	GASSTEAN	O2	RAWPROD	RECYCLE	SYNGAS
Temperat	339	783	800	533	1081	823	305	299	475	305	373	375	368	308	308	781	313	313	781	781	305	601	298		313	313	783	458	533	313	1100
Pressure	27	1	38	20	122	38	0	38	38	0	1	20	1	0	0	38	20	20	38	38	1	1	1		20	20	1	29	23	20	1
Mass VFra	1	1	1	1	1	1	0	0	0	1	1	0	0	1	0	1	1	1	1	1	0	1	0		1	0	1	1	1	1	1
Mass SFra	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
Mass Flow	63625	117247	456476	41868	398301	40287	398301	339229	339229	398301	419738	229903	2097	185	1912	4163	300	249	811	12914	398301	377871	339229		41868	18011	88697	22718	229903	63625	254800
Volume Fl	35950	7062220	719503	50088	267868	67127	6721	5694	7061	191746000	7372670	96788	39	77642	32	6584	211	175	1283	20427	6721	10796300	5687		29439	415	5342550	15339	300875	44738	18836000
Mass Flow																															
CO	37382	0	0	24599	0	0	0	0	0	0	330	62304	0	0	0	0	176	146	0	0	0	0	0		24599	0	0	0	62304	37382	90483
H2	0	0	0	0	0	0	0	0	0	0	3	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	5844
CO2	9266	0	0	6097	0	0	0	0	0	0	77218	15443	0	0	0	0	44	36	0	0	0	0	0		6097	0	0	0	15443	9266	81716
H2O	0	117247	456476	0	398301	40287	398301	339229	339229	398301	19350	100531	2097	184	1912	4163	0	0	811	12914	398301	0	339229		0	0	88697	0	100531	0	60604
CH4	4878	0	0	3210	0	0	0	0	0	0	0	13448	0	0	0	0	23	19	0	0	0	0	0		3210	0	0	0	13448	4878	16153
C4H8	971	0	0	639	0	0	0	0	0	0	0	1619	0	0	0	0	5	4	0	0	0	0	0		639	0	0	0	1619	971	0
C4H10	11128	0	0	7323	0	0	0	0	0	0	0	18547	0	0	0	0	52	44	0	0	0	0	0		7323	0	0	0	18547	11128	0
C9H18	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C9H20	0	0	0	0	0	0	0	0	0	0	0	7017	0	0	0	0	0	0	0	0	0	0	0		0	7017	0	0	7017	0	0
C15H30	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
C15H32	0	0	0	0	0	0	0	0	0	0	0	10995	0	0	0	0	0	0	0	0	0	0	0		0	10995	0	0	10995	0	0
C21H44	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
MEOH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	298518	0	0	0	0	0	0	0	0	0	0	298518	0		0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	24319	0	0	0	0	0	0	0	0	0	0	79353	0		0	0	0	22718	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
NAPHT-0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
MTBE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
N-HEX-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
TRIMEPE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
ETHYLBEI	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
CUMENE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
CRESOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
BIPHENY	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
PHENOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
BENZENE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
TOLUENE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
XYLENE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
STYRENE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0

APPENDIX 2: ENERGY USAGE AND PINCH POINT ANALYSIS

APPENDIX 2.1

Table A 1: First Order Model of Refinery Section of GFT Process



Component			Naphtha Hydrotreating	Kerosene Hydrotreating	Heavy Distillate Hydrotreating	Naphtha Reforming	Catalytic Cracking	Isomerisation
Product Distribution (vol%of unit feed)								
Hydrogen						8.76	1.04	
Methane			0.06	0.22	0.23	0.29	1.24	
Ethane			0.09	0.01	0.01	0.68	0.79	
Ethalene							1.41	
Propane			0.06	0.03	0.04	2.23		
N-Butane			0.35			1.07		
Iso-Butane						0.56		
C3 mixture							13.02	
C4 mixture							14.48	
Light Gases			0.56	0.26	0.28	13.59	31.98	0.54
Naphtha			99.64					
Light Naphtha			49.82				14.44	
Heavy Naphtha			49.82				10.10	
Isomerate								99.49
Reformate						87.50		
Gasoline Components			99.64	0.00	0.00	87.50	24.54	
Waxes					100.60			
Distillate				100.08				
Diesel							19.38	
Aromatics							38.60	
Olefins							4.70	
Diesel Components			0.00	100.08	0.00	0.00	62.68	
Utilities	**Units**		**Usage Per Litre of Unit Feed**					
Fuel Gas	Mass	kg	3.48E-03	3.88E-03	4.15E-03	3.75E-03	1.81E-02	0.00E+00
Electricity	Electricity	kWhr	4.28E-03	1.02E-02	1.11E-02	3.05E-02	0.00E+00	1.30E-02
Steam	Mass	kg	5.71E-03	1.71E-02	2.00E-02	1.65E-01	1.08E-01	4.95E-01
Cooling Water	Volume	l	2.38E-02	7.14E-02	7.14E-02	2.17E+00	2.86E-01	2.19E-01
Catalyst Consumption	Mass	kg/ton	3.55E-05	3.93E-05		9.14E-03	2.26E-03	1.46E-07
Air Emmisions			**kg Emission Per litre of Unit Feed**					
Carbon Dioxide			9.57E-03	1.07E-02	1.14E-02	1.03E-02	4.97E-02	0.00E+00
Steam			7.83E-03	8.73E-03	9.33E-03	8.43E-03	4.06E-02	0.00E+00
NOx			3.45E-06	1.35E-05	1.92E-05	4.87E-06	4.25E-08	0.00E+00
Carbon Monoxide			5.31E-07	0.00E+00	0.00E+00	4.87E-06	9.71E-06	0.00E+00
Particulates			1.06E-06	0.00E+00	1.07E-06	2.12E-07	9.03E-06	0.00E+00
Hydrocarbons			0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.57E-07	0.00E+00
Aldehydes			0.00E+00	0.00E+00	0.00E+00	0.00E+00	1.35E-08	0.00E+00
Water Effluent (liquid Phase)			1.30E-03	1.44E-03		2.32E-03	1.08E-01	

APPENDIX 2.2: ENERGY USAGE OF MAJOR PROCESSING UNITS

Table A 2: SHF Process Energy Consumption

Block	Unit	Refrigeration	Electricity	Heat	Cooling	Cooling Water
		kW/MW input	kW/MW input	kW/MW input	kW/MW input	Kg/hr/MW input
	SO2 Burner				-8.59E+00	
	Pre-Steam Heat			7.64E+00		
	Pre-Treatment			1.65E+00	-1.75E+01	
	Water Pump		1.83E-03			
	Pre-Steam Steam			1.54E+01		
Block 2	An-Aerobic Batch Cultivatio	-1.21E+01	4.02E+00			
Block 3	Sacharification				-3.43E+01	
	Fermention	-1.45E+01	4.85E+00			
Bock 4	CO2 F1			4.26E-01		
	CO2 F2				-1.80E+00	
	Beer Column			1.54E+02	-3.01E+01	
	Rectifier			1.88E+01	-3.23E+01	
	Molecular Siev	-1.12E+01	3.73E+00	8.12E-01	-5.87E+00	
Block 5	Evaporation			4.99E+01		
	Pneumatic Press		1.59E+00		6.45E+00	
Block 7	De-Air			2.25E+01		

Table A 3: SSF Process Energy Usage

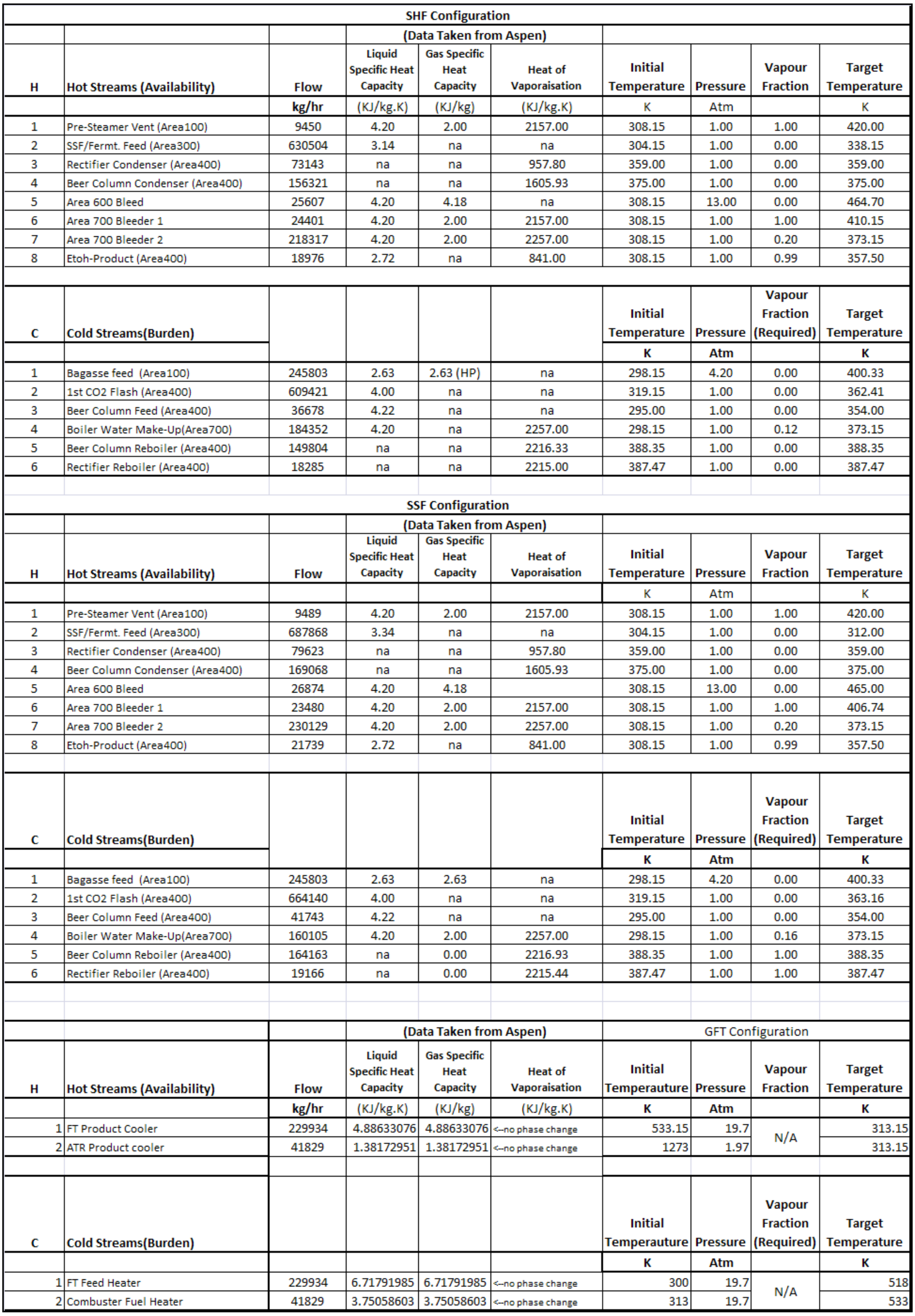
Block	Unit	Refrigeration	Electricity	Heat	Cooling	Cooling Water
		kW/MW input	kW/MW input	kW/MW input	kW/MW input	Kg/hr/MW input
	SO2 Burner				-8.59E+00	
	Pre-Steam Heat			7.10E+00		
	Pre-Treatment		1.60E+00	1.70E+00	-1.75E+01	
	Water Pump		1.63E-03			
	Pre-Steam Steam			1.58E+01		
Block 2	An-Aerobic Batch Cultivatio	-7.17E-01	2.39E-01			
	Aerobic Batch Cultivation	-1.41E+01	4.70E+00			
Block 3	SSF Reactor	3.12E+01	1.40E+01			
		3.12E+01	5.04E+01		-4.26E+00	3.93E+02
Bock 4	CO2 F1			1.81E-01		
	CO2 F2				-2.10E+00	
	Beer Column			1.68E+02	-3.85E+01	
	Rectifier			1.97E+01	-3.53E+01	
	Molecular Siev	-1.28E+01	4.28E+00	9.32E-01	-6.80E+00	
Block 5	Evaporation			5.13E+01		
	Pneumatic Press		1.59E+00		-6.93E+00	
Block 7	De-Air			2.71E+01		

Table A 4: GFT Process Energy Consumption

Unit	Refrigeration	Electricity	Heat	Fuel	Cooling	Cooling Water
	kW/MW input	kW/MW input	kW/MW input	kW/MW input	kW/MW input	Kg/hr/MW input
Drying and Handling		1.12E+01	2.87E+01			
Oxygen Production		3.87E+01				
Gasification			1.39E+02			
Rectisol		2.38E+00	2.33E+01			
Refrigeration	-1.00E+01	3.35E+00				
FT Compressor		9.72E+00				
FT Reactor Cooler					-2.25E+01	
ATR		1.07E+00				
Distillation		1.99E-01	2.05E+00	6.19E-01		3.21E+02
Refinery		4.60E-01	9.24E+00	6.64E-01		2.91E+01

APPENDIX 2.3: PINCH POINT ANALYSES

Table A 5: Stream Data



SHF Configuration									
			(Data Taken from Aspen)						
H	Hot Streams (Availability)	Flow	Liquid Specific Heat Capacity	Gas Specific Heat Capacity	Heat of Vaporaisation	Initial Temperature	Pressure	Vapour Fraction	Target Temperature
		kg/hr	(KJ/kg.K)	(KJ/kg)	(KJ/kg.K)	K	Atm		K
1	Pre-Steamer Vent (Area100)	9450	4.20	2.00	2157.00	308.15	1.00	1.00	420.00
2	SSF/Fermt. Feed (Area300)	630504	3.14	na	na	304.15	1.00	0.00	338.15
3	Rectifier Condenser (Area400)	73143	na	na	957.80	359.00	1.00	0.00	359.00
4	Beer Column Condenser (Area400)	156321	na	na	1605.93	375.00	1.00	0.00	375.00
5	Area 600 Bleed	25607	4.20	4.18	na	308.15	13.00	0.00	464.70
6	Area 700 Bleeder 1	24401	4.20	2.00	2157.00	308.15	1.00	1.00	410.15
7	Area 700 Bleeder 2	218317	4.20	2.00	2257.00	308.15	1.00	0.20	373.15
8	Etoh-Product (Area400)	18976	2.72	na	841.00	308.15	1.00	0.99	357.50
C	Cold Streams(Burden)					Initial Temperature	Pressure	Vapour Fraction (Required)	Target Temperature
						K	Atm		K
1	Bagasse feed (Area100)	245803	2.63	2.63 (HP)	na	298.15	4.20	0.00	400.33
2	1st CO2 Flash (Area400)	609421	4.00	na	na	319.15	1.00	0.00	362.41
3	Beer Column Feed (Area400)	36678	4.22	na	na	295.00	1.00	0.00	354.00
4	Boiler Water Make-Up(Area700)	184352	4.20	na	2257.00	298.15	1.00	0.12	373.15
5	Beer Column Reboiler (Area400)	149804	na	na	2216.33	388.35	1.00	0.00	388.35
6	Rectifier Reboiler (Area400)	18285	na	na	2215.00	387.47	1.00	0.00	387.47
SSF Configuration									
			(Data Taken from Aspen)						
H	Hot Streams (Availability)	Flow	Liquid Specific Heat Capacity	Gas Specific Heat Capacity	Heat of Vaporaisation	Initial Temperature	Pressure	Vapour Fraction	Target Temperature
						K	Atm		K
1	Pre-Steamer Vent (Area100)	9489	4.20	2.00	2157.00	308.15	1.00	1.00	420.00
2	SSF/Fermt. Feed (Area300)	687868	3.34	na	na	304.15	1.00	0.00	312.00
3	Rectifier Condenser (Area400)	79623	na	na	957.80	359.00	1.00	0.00	359.00
4	Beer Column Condenser (Area400)	169068	na	na	1605.93	375.00	1.00	0.00	375.00
5	Area 600 Bleed	26874	4.20	4.18		308.15	13.00	0.00	465.00
6	Area 700 Bleeder 1	23480	4.20	2.00	2157.00	308.15	1.00	1.00	406.74
7	Area 700 Bleeder 2	230129	4.20	2.00	2257.00	308.15	1.00	0.20	373.15
8	Etoh-Product (Area400)	21739	2.72	na	841.00	308.15	1.00	0.99	357.50
C	Cold Streams(Burden)					Initial Temperature	Pressure	Vapour Fraction (Required)	Target Temperature
						K	Atm		K
1	Bagasse feed (Area100)	245803	2.63	2.63	na	298.15	4.20	0.00	400.33
2	1st CO2 Flash (Area400)	664140	4.00	na	na	319.15	1.00	0.00	363.16
3	Beer Column Feed (Area400)	41743	4.22	na	na	295.00	1.00	0.00	354.00
4	Boiler Water Make-Up(Area700)	160105	4.20	2.00	2257.00	298.15	1.00	0.16	373.15
5	Beer Column Reboiler (Area400)	164163	na	0.00	2216.93	388.35	1.00	1.00	388.35
6	Rectifier Reboiler (Area400)	19166	na	0.00	2215.44	387.47	1.00	1.00	387.47
			(Data Taken from Aspen)			GFT Configuration			
H	Hot Streams (Availability)	Flow	Liquid Specific Heat Capacity	Gas Specific Heat Capacity	Heat of Vaporaisation	Initial Temperauture	Pressure	Vapour Fraction	Target Temperature
		kg/hr	(KJ/kg.K)	(KJ/kg)	(KJ/kg.K)	K	Atm		K
1	FT Product Cooler	229934	4.88633076	4.88633076	<--no phase change	533.15	19.7	N/A	313.15
2	ATR Product cooler	41829	1.38172951	1.38172951	<--no phase change	1273	1.97		313.15
C	Cold Streams(Burden)					Initial Temperauture	Pressure	Vapour Fraction (Required)	Target Temperature
						K	Atm		K
1	FT Feed Heater	229934	6.71791985	6.71791985	<--no phase change	300	19.7	N/A	518
2	Combuster Fuel Heater	41829	3.75058603	3.75058603	<--no phase change	313	19.7		533

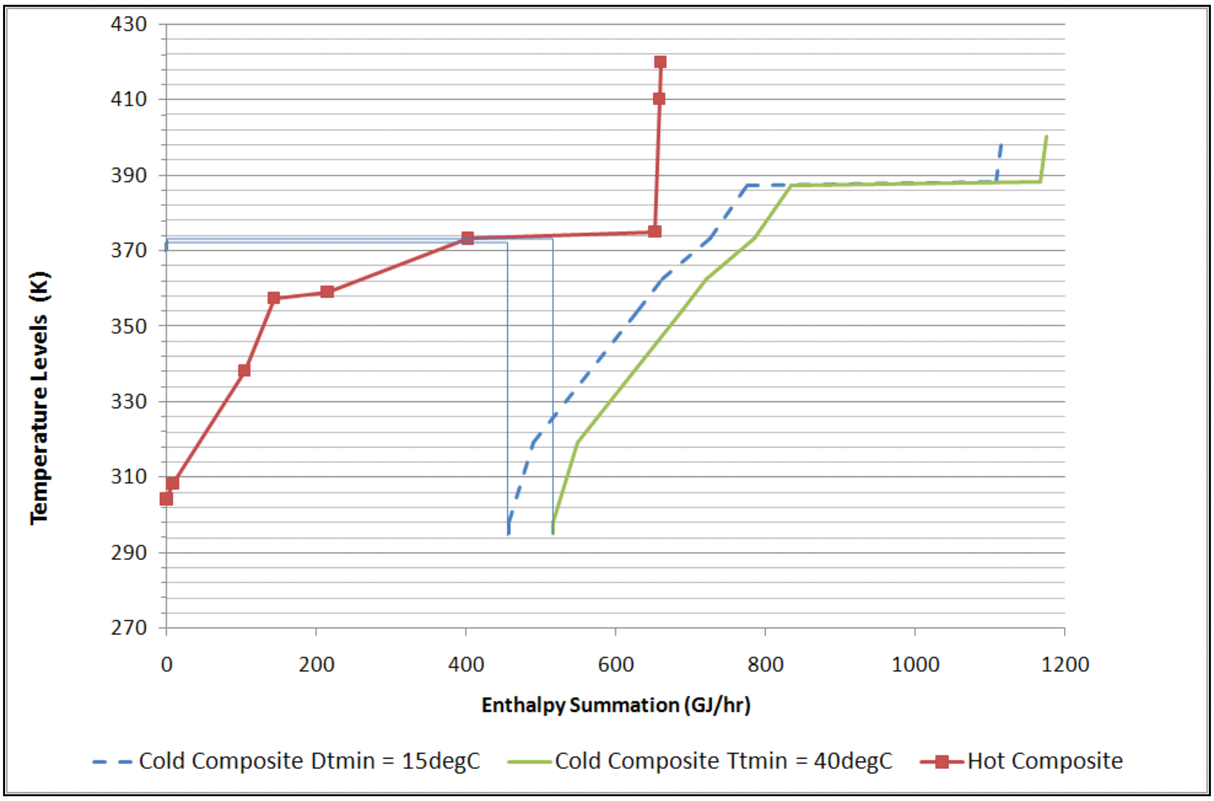
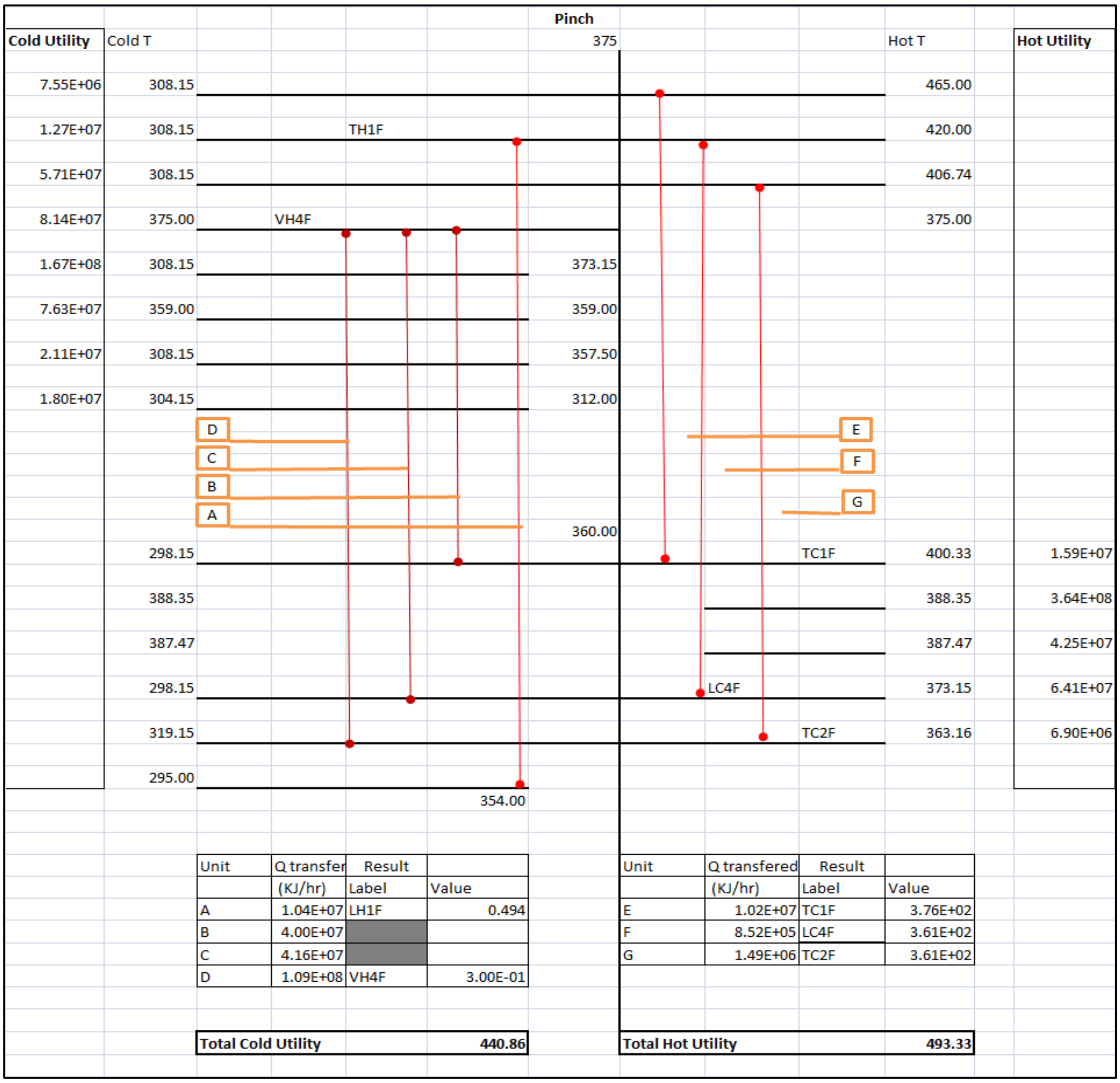


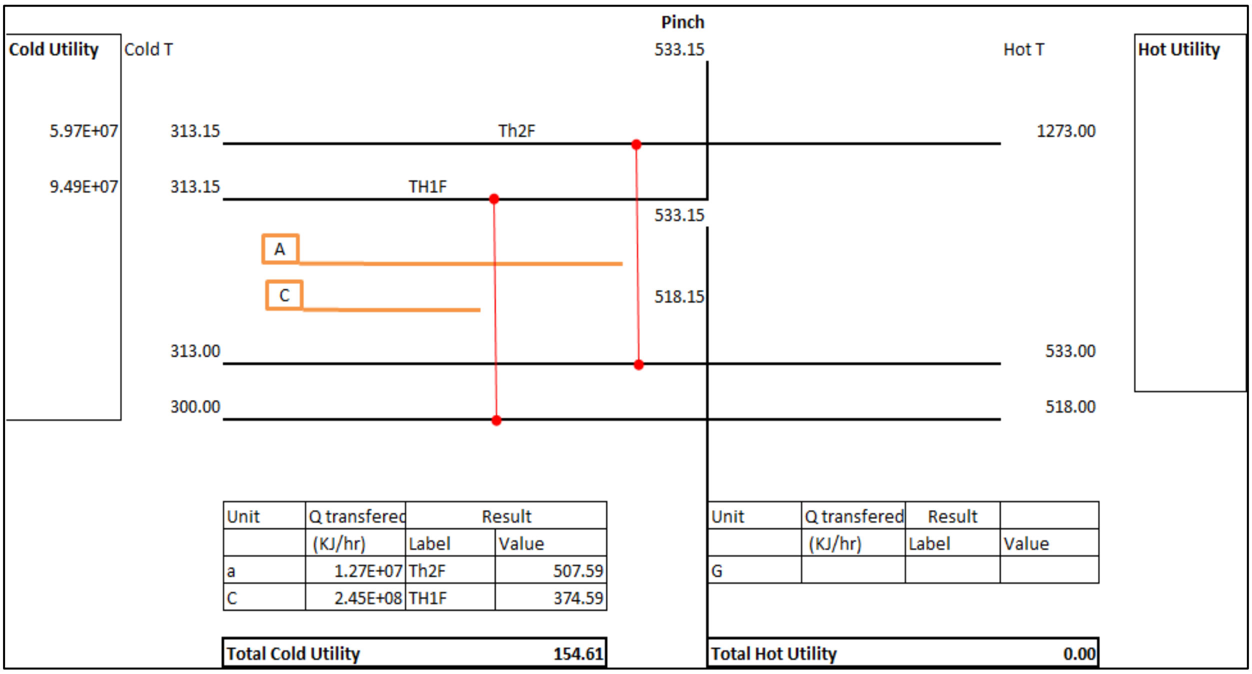
Figure A 1: SHF Composite Curves



Unit	Q transfer	Result	
	(KJ/hr)	Label	Value
A	1.04E+07	LH1F	0.494
B	4.00E+07	▨	
C	4.16E+07	▨	
D	1.09E+08	VH4F	3.00E-01
Total Cold Utility			**440.86**

Unit	Q transfered	Result	
	(KJ/hr)	Label	Value
E	1.02E+07	TC1F	3.76E+02
F	8.52E+05	LC4F	3.61E+02
G	1.49E+06	TC2F	3.61E+02
Total Hot Utility			**493.33**

Figure A 2: Heat Exchange Network for SSF



Unit	Q transfered	Result	
	(KJ/hr)	Label	Value
a	1.27E+07	Th2F	507.59
C	2.45E+08	TH1F	374.59

Total Cold Utility	154.61

Unit	Q transfered	Result	
	(KJ/hr)	Label	Value
G			

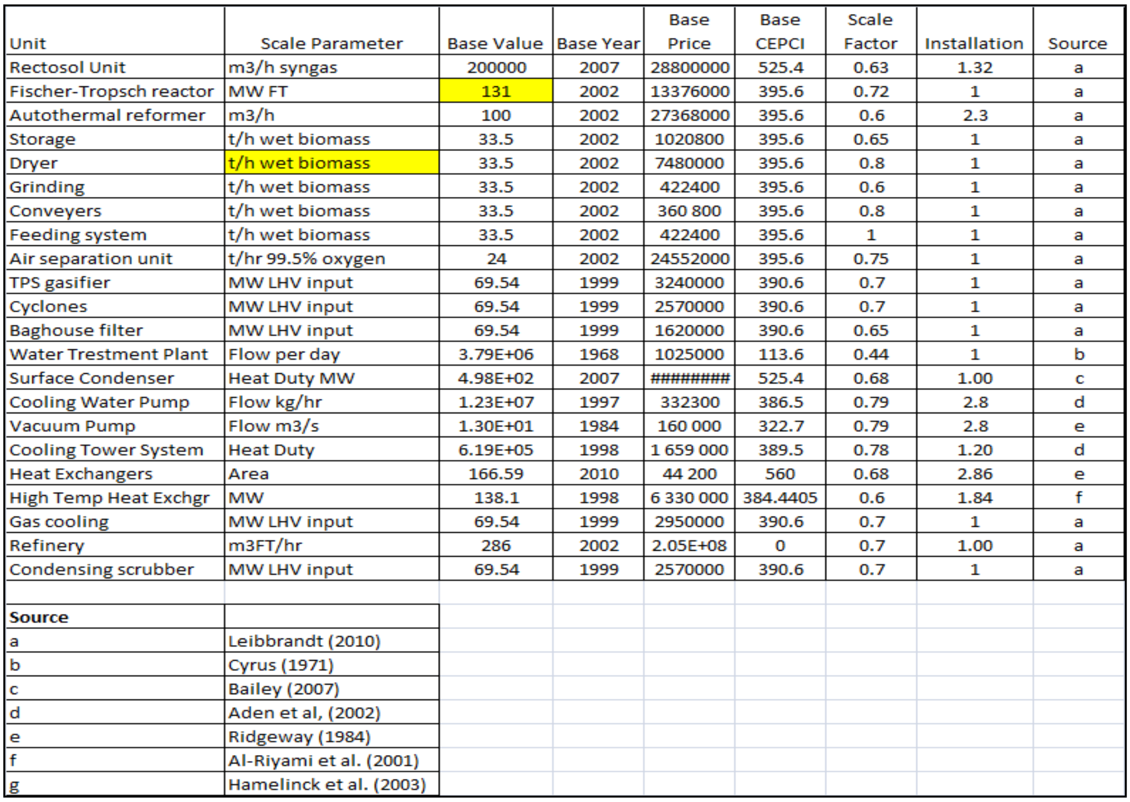
Total Hot Utility	0.00

Figure A 3: Heat Exchange Network for GFT

APPENDIX 3: ECONOMICS

APPENDIX 3.1 CAPITAL COSTS ESTIMATION FROM LITERATURE

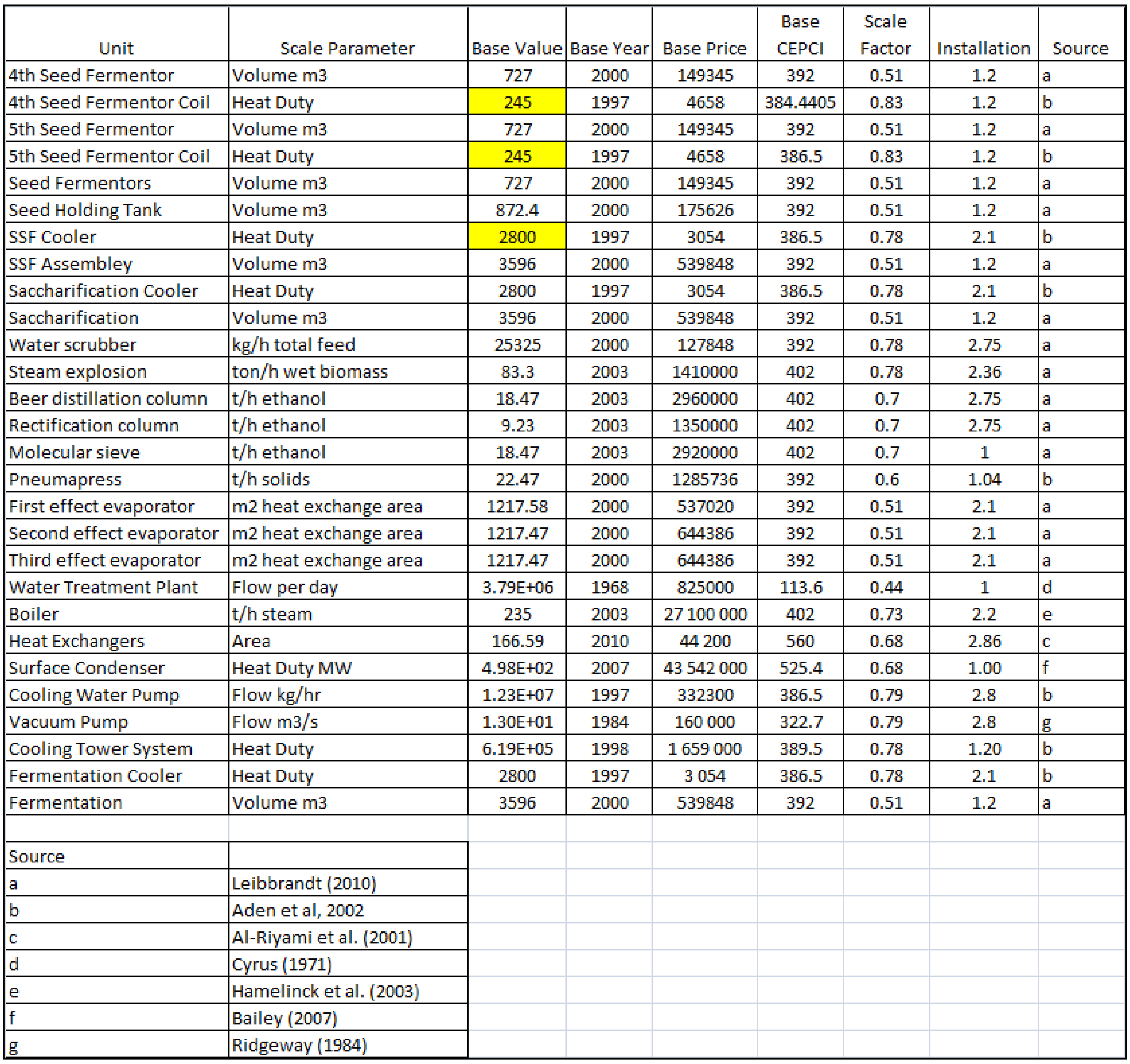
Table A 6: GFT Scenario



Unit	Scale Parameter	Base Value	Base Year	Base Price	Base CEPCI	Scale Factor	Installation	Source
Rectosol Unit	m3/h syngas	200000	2007	28800000	525.4	0.63	1.32	a
Fischer-Tropsch reactor	MW FT	131	2002	13376000	395.6	0.72	1	a
Autothermal reformer	m3/h	100	2002	27368000	395.6	0.6	2.3	a
Storage	t/h wet biomass	33.5	2002	1020800	395.6	0.65	1	a
Dryer	t/h wet biomass	33.5	2002	7480000	395.6	0.8	1	a
Grinding	t/h wet biomass	33.5	2002	422400	395.6	0.6	1	a
Conveyers	t/h wet biomass	33.5	2002	360 800	395.6	0.8	1	a
Feeding system	t/h wet biomass	33.5	2002	422400	395.6	1	1	a
Air separation unit	t/hr 99.5% oxygen	24	2002	24552000	395.6	0.75	1	a
TPS gasifier	MW LHV input	69.54	1999	3240000	390.6	0.7	1	a
Cyclones	MW LHV input	69.54	1999	2570000	390.6	0.7	1	a
Baghouse filter	MW LHV input	69.54	1999	1620000	390.6	0.65	1	a
Water Trestment Plant	Flow per day	3.79E+06	1968	1025000	113.6	0.44	1	b
Surface Condenser	Heat Duty MW	4.98E+02	2007	#########	525.4	0.68	1.00	c
Cooling Water Pump	Flow kg/hr	1.23E+07	1997	332300	386.5	0.79	2.8	d
Vacuum Pump	Flow m3/s	1.30E+01	1984	160 000	322.7	0.79	2.8	e
Cooling Tower System	Heat Duty	6.19E+05	1998	1 659 000	389.5	0.78	1.20	d
Heat Exchangers	Area	166.59	2010	44 200	560	0.68	2.86	e
High Temp Heat Exchgr	MW	138.1	1998	6 330 000	384.4405	0.6	1.84	f
Gas cooling	MW LHV input	69.54	1999	2950000	390.6	0.7	1	a
Refinery	m3FT/hr	286	2002	2.05E+08	0	0.7	1.00	a
Condensing scrubber	MW LHV input	69.54	1999	2570000	390.6	0.7	1	a

Source	
a	Leibbrandt (2010)
b	Cyrus (1971)
c	Bailey (2007)
d	Aden et al, (2002)
e	Ridgeway (1984)
f	Al-Riyami et al. (2001)
g	Hamelinck et al. (2003)

Table A 7: Biological Scenarios



Unit	Scale Parameter	Base Value	Base Year	Base Price	Base CEPCI	Scale Factor	Installation	Source
4th Seed Fermentor	Volume m3	727	2000	149345	392	0.51	1.2	a
4th Seed Fermentor Coil	Heat Duty	245	1997	4658	384.4405	0.83	1.2	b
5th Seed Fermentor	Volume m3	727	2000	149345	392	0.51	1.2	a
5th Seed Fermentor Coil	Heat Duty	245	1997	4658	386.5	0.83	1.2	b
Seed Fermentors	Volume m3	727	2000	149345	392	0.51	1.2	a
Seed Holding Tank	Volume m3	872.4	2000	175626	392	0.51	1.2	a
SSF Cooler	Heat Duty	2800	1997	3054	386.5	0.78	2.1	b
SSF Assembley	Volume m3	3596	2000	539848	392	0.51	1.2	a
Saccharification Cooler	Heat Duty	2800	1997	3054	386.5	0.78	2.1	b
Saccharification	Volume m3	3596	2000	539848	392	0.51	1.2	a
Water scrubber	kg/h total feed	25325	2000	127848	392	0.78	2.75	a
Steam explosion	ton/h wet biomass	83.3	2003	1410000	402	0.78	2.36	a
Beer distillation column	t/h ethanol	18.47	2003	2960000	402	0.7	2.75	a
Rectification column	t/h ethanol	9.23	2003	1350000	402	0.7	2.75	a
Molecular sieve	t/h ethanol	18.47	2003	2920000	402	0.7	1	a
Pneumapress	t/h solids	22.47	2000	1285736	392	0.6	1.04	b
First effect evaporator	m2 heat exchange area	1217.58	2000	537020	392	0.51	2.1	a
Second effect evaporator	m2 heat exchange area	1217.47	2000	644386	392	0.51	2.1	a
Third effect evaporator	m2 heat exchange area	1217.47	2000	644386	392	0.51	2.1	a
Water Treatment Plant	Flow per day	3.79E+06	1968	825000	113.6	0.44	1	d
Boiler	t/h steam	235	2003	27 100 000	402	0.73	2.2	e
Heat Exchangers	Area	166.59	2010	44 200	560	0.68	2.86	c
Surface Condenser	Heat Duty MW	4.98E+02	2007	43 542 000	525.4	0.68	1.00	f
Cooling Water Pump	Flow kg/hr	1.23E+07	1997	332300	386.5	0.79	2.8	b
Vacuum Pump	Flow m3/s	1.30E+01	1984	160 000	322.7	0.79	2.8	g
Cooling Tower System	Heat Duty	6.19E+05	1998	1 659 000	389.5	0.78	1.20	b
Fermentation Cooler	Heat Duty	2800	1997	3 054	386.5	0.78	2.1	b
Fermentation	Volume m3	3596	2000	539848	392	0.51	1.2	a

Source	
a	Leibbrandt (2010)
b	Aden et al, 2002
c	Al-Riyami et al. (2001)
d	Cyrus (1971)
e	Hamelinck et al. (2003)
f	Bailey (2007)
g	Ridgeway (1984)

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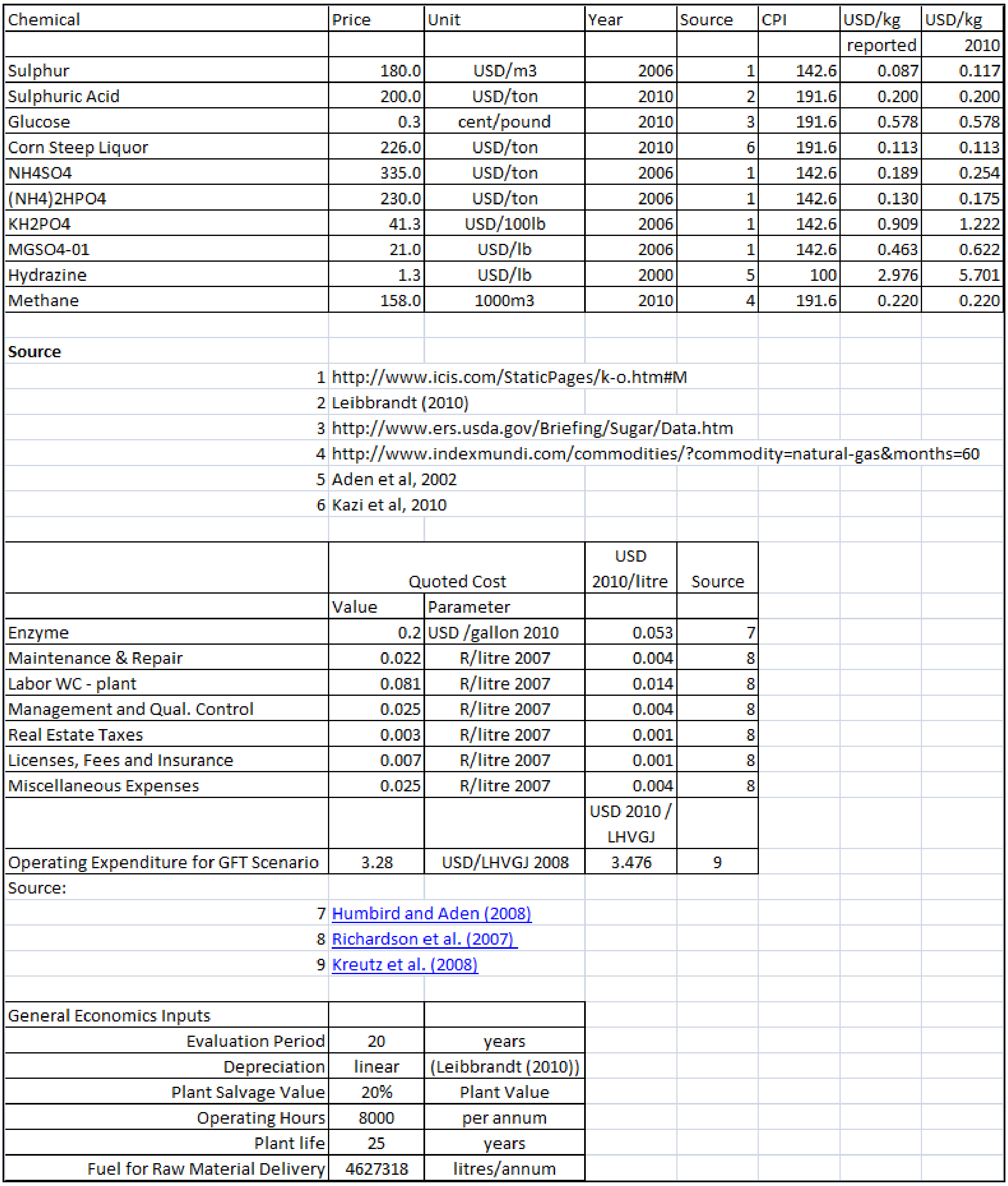
APPENDIX 3.2: NON STOCHASTIC AND GENERAL ECONOMIC INPUTS

Table A 8: Non Stochastic and General Economic inputs



Chemical	Price	Unit	Year	Source	CPI	USD/kg reported	USD/kg 2010
Sulphur	180.0	USD/m3	2006	1	142.6	0.087	0.117
Sulphuric Acid	200.0	USD/ton	2010	2	191.6	0.200	0.200
Glucose	0.3	cent/pound	2010	3	191.6	0.578	0.578
Corn Steep Liquor	226.0	USD/ton	2010	6	191.6	0.113	0.113
NH4SO4	335.0	USD/ton	2006	1	142.6	0.189	0.254
(NH4)2HPO4	230.0	USD/ton	2006	1	142.6	0.130	0.175
KH2PO4	41.3	USD/100lb	2006	1	142.6	0.909	1.222
MGSO4-01	21.0	USD/lb	2006	1	142.6	0.463	0.622
Hydrazine	1.3	USD/lb	2000	5	100	2.976	5.701
Methane	158.0	1000m3	2010	4	191.6	0.220	0.220

Source

1 http://www.icis.com/StaticPages/k-o.htm#M
2 Leibbrandt (2010)
3 http://www.ers.usda.gov/Briefing/Sugar/Data.htm
4 http://www.indexmundi.com/commodities/?commodity=natural-gas&months=60
5 Aden et al, 2002
6 Kazi et al, 2010

	Quoted Cost Value	Quoted Cost Parameter	USD 2010/litre	Source
Enzyme	0.2	USD /gallon 2010	0.053	7
Maintenance & Repair	0.022	R/litre 2007	0.004	8
Labor WC - plant	0.081	R/litre 2007	0.014	8
Management and Qual. Control	0.025	R/litre 2007	0.004	8
Real Estate Taxes	0.003	R/litre 2007	0.001	8
Licenses, Fees and Insurance	0.007	R/litre 2007	0.001	8
Miscellaneous Expenses	0.025	R/litre 2007	0.004	8
			USD 2010 / LHVGJ	
Operating Expenditure for GFT Scenario	3.28	USD/LHVGJ 2008	3.476	9

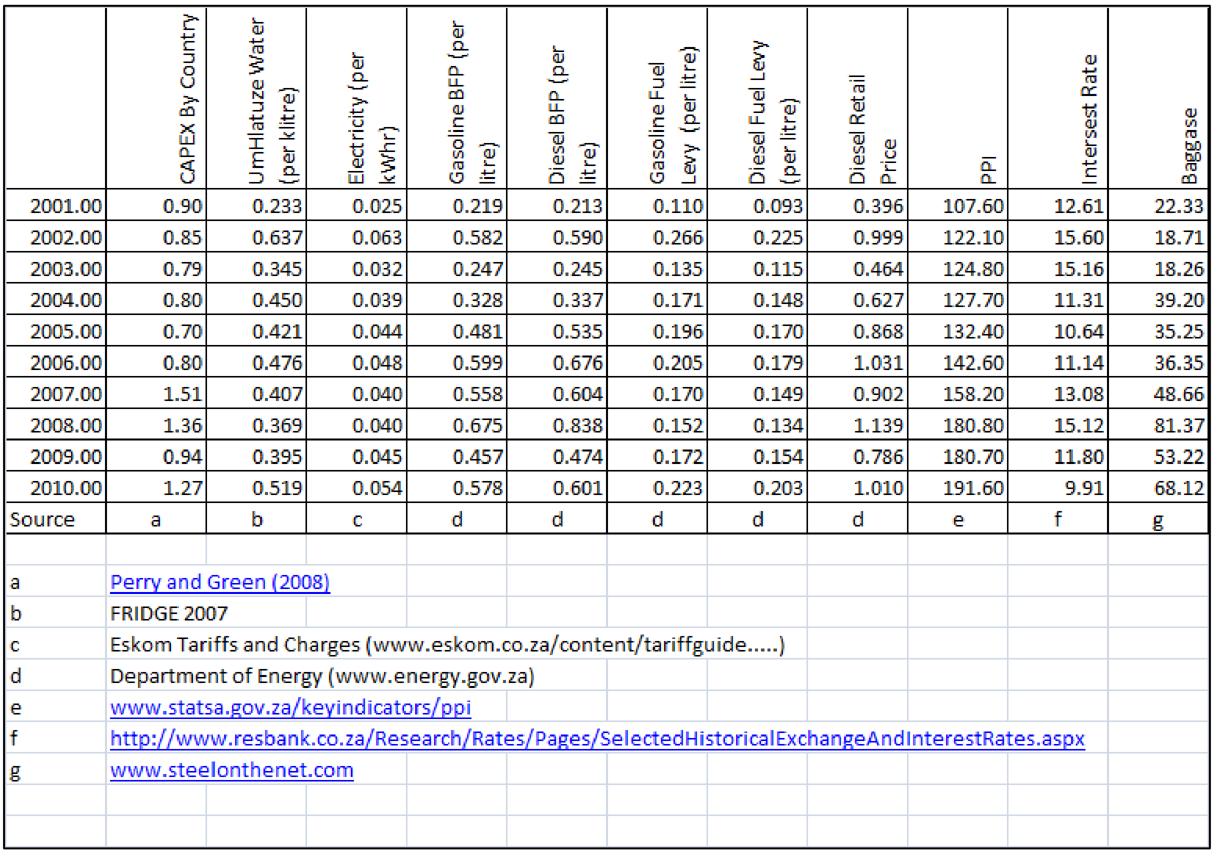
Source:

7 Humbird and Aden (2008)
8 Richardson et al. (2007)
9 Kreutz et al. (2008)

General Economics Inputs		
Evaluation Period	20	years
Depreciation	linear	(Leibbrandt (2010))
Plant Salvage Value	20%	Plant Value
Operating Hours	8000	per annum
Plant life	25	years
Fuel for Raw Material Delivery	4627318	litres/annum

APPENDIX 3.3: STOCHASTIC INPUT FOR MONTE CARLO SIMULATION

Table A 9: Stochastic Input for Monte Carlo Simulation



	CAPEX By Country	UmHlatuze Water (per klitre)	Electricity (per kWhr)	Gasoline BFP (per litre)	Diesel BFP (per litre)	Gasoline Fuel Levy (per litre)	Diesel Fuel Levy (per litre)	Diesel Retail Price	PPI	Intersest Rate	Baggase
2001.00	0.90	0.233	0.025	0.219	0.213	0.110	0.093	0.396	107.60	12.61	22.33
2002.00	0.85	0.637	0.063	0.582	0.590	0.266	0.225	0.999	122.10	15.60	18.71
2003.00	0.79	0.345	0.032	0.247	0.245	0.135	0.115	0.464	124.80	15.16	18.26
2004.00	0.80	0.450	0.039	0.328	0.337	0.171	0.148	0.627	127.70	11.31	39.20
2005.00	0.70	0.421	0.044	0.481	0.535	0.196	0.170	0.868	132.40	10.64	35.25
2006.00	0.80	0.476	0.048	0.599	0.676	0.205	0.179	1.031	142.60	11.14	36.35
2007.00	1.51	0.407	0.040	0.558	0.604	0.170	0.149	0.902	158.20	13.08	48.66
2008.00	1.36	0.369	0.040	0.675	0.838	0.152	0.134	1.139	180.80	15.12	81.37
2009.00	0.94	0.395	0.045	0.457	0.474	0.172	0.154	0.786	180.70	11.80	53.22
2010.00	1.27	0.519	0.054	0.578	0.601	0.223	0.203	1.010	191.60	9.91	68.12
Source	a	b	c	d	d	d	d	d	e	f	g

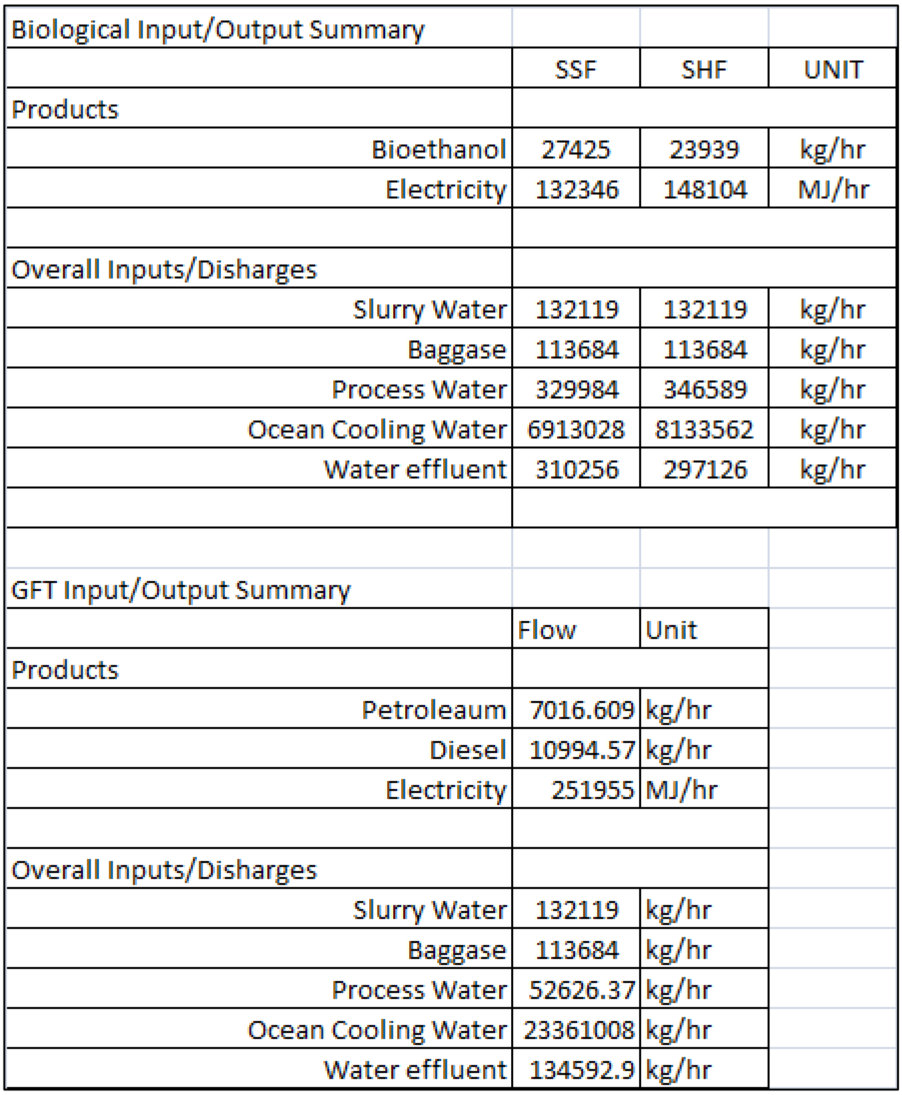
a	Perry and Green (2008)
b	FRIDGE 2007
c	Eskom Tariffs and Charges (www.eskom.co.za/content/tariffguide.....)
d	Department of Energy (www.energy.gov.za)
e	www.statsa.gov.za/keyindicators/ppi
f	http://www.resbank.co.za/Research/Rates/Pages/SelectedHistoricalExchangeAndInterestRates.aspx
g	www.steelonthenet.com

NB: www.steelonthenet.com is a consultancy for major steel companies. They publish free relevant data for commodities used by the steel industry from databanks that are very expensive. Price of bagasse was calculated on the energy equivalence with coal. The prices given on the databanks from www.steelonthenet.com was in agreement with the spot prices published by the International Energy Agency.

APPENDIX 4: LIFE CYCLE ASSESSMENTS: LIFE CYCLE INVENTORY

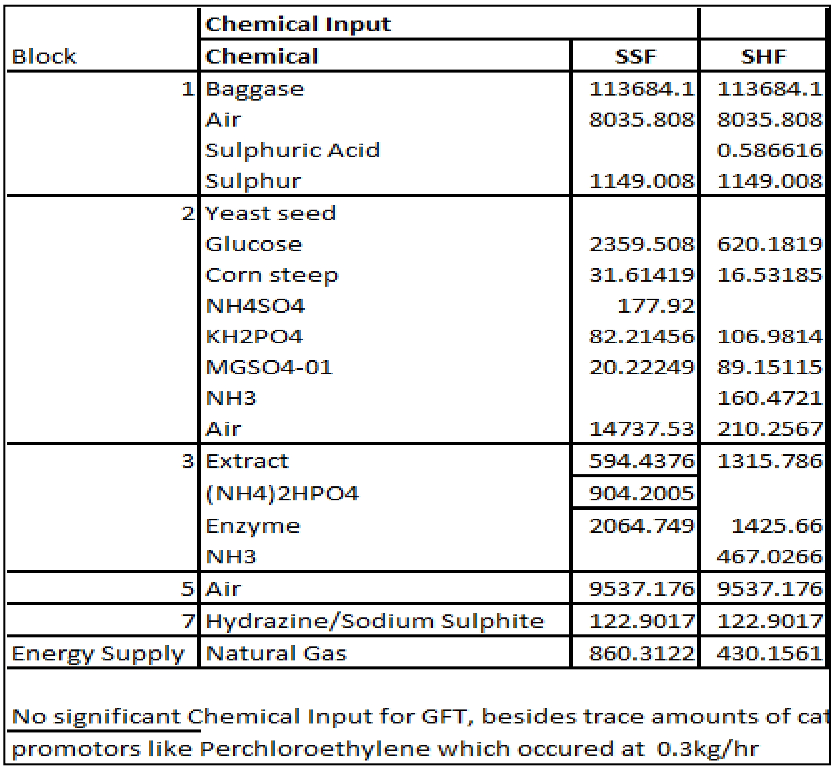
APPENDIX 4.1: PROCESS INPUTS

Table A 10: Process Inputs for LCI



Biological Input/Output Summary			
	SSF	SHF	UNIT
Products			
Bioethanol	27425	23939	kg/hr
Electricity	132346	148104	MJ/hr
Overall Inputs/Disharges			
Slurry Water	132119	132119	kg/hr
Baggase	113684	113684	kg/hr
Process Water	329984	346589	kg/hr
Ocean Cooling Water	6913028	8133562	kg/hr
Water effluent	310256	297126	kg/hr
GFT Input/Output Summary			
	Flow	Unit	
Products			
Petroleaum	7016.609	kg/hr	
Diesel	10994.57	kg/hr	
Electricity	251955	MJ/hr	
Overall Inputs/Disharges			
Slurry Water	132119	kg/hr	
Baggase	113684	kg/hr	
Process Water	52626.37	kg/hr	
Ocean Cooling Water	23361008	kg/hr	
Water effluent	134592.9	kg/hr	

Table A 11: Chemical Inputs



	Chemical Input		
Block	Chemical	SSF	SHF
1	Baggase	113684.1	113684.1
	Air	8035.808	8035.808
	Sulphuric Acid		0.586616
	Sulphur	1149.008	1149.008
2	Yeast seed		
	Glucose	2359.508	620.1819
	Corn steep	31.61419	16.53185
	NH4SO4	177.92	
	KH2PO4	82.21456	106.9814
	MGSO4-01	20.22249	89.15115
	NH3		160.4721
	Air	14737.53	210.2567
3	Extract	594.4376	1315.786
	(NH4)2HPO4	904.2005	
	Enzyme	2064.749	1425.66
	NH3		467.0266
5	Air	9537.176	9537.176
7	Hydrazine/Sodium Sulphite	122.9017	122.9017
Energy Supply	Natural Gas	860.3122	430.1561

No significant Chemical Input for GFT, besides trace amounts of cat
promotors like Perchloroethylene which occured at 0.3kg/hr

For all processes, a truck that drove 15424394.7 km/yr was included for raw material delivery.

APPENDIX 4.2: EMISSIONS

Table A 12: SHF Process

	Block4	Block5	Block6	Cooling Tower
Ethanol	3.08E-01	3.03E-01	2.21E-02	3.94E+01
Carbon Dioxide	2.28E+04		1.36E+05	
Oxygen	2.85E+01	2.22E+03	7.90E+03	
Nitrogen		4.20E+05	3.33E+05	
Steam	1.49E+02	4.77E+02	1.91E+05	8.19E+01
Acetic Acid			1.47E+01	1.54E+03
Sulphur Dioxide			3.10E+01	7.21E-01
Furfural			3.64E+00	2.14E+00
Hydrazine			3.85E+00	
Xylitol			4.77E+00	
Methane			4.30E+00	
Nitrogen Dioxide			4.60E-02	
Carbon Monoxide			2.80E-03	
Sulphuric Acid			5.70E-01	
Water Effluent	297125.6	kg/hr		

Table A 13: SSF Process

	Air Emission Data			
Constituent	Emission Rate (kg/hr)			
	Block4	Block5	Block6	Cooling Tower
Ethanol	3.61E-01	3.38E-01	2.66E-02	4.48E+01
Carbon Dioxide	2.58E+04		1.41E+05	
Oxygen	6.20E+01	2.22E+03	2.39E+03	
Nitrogen	2.30E-01	7.31E+03	4.20E+05	
Steam	1.69E+02	4.89E+02	2.23E+05	2.94E+04
Acetic Acid			1.57E+01	7.71E+01
Sulphur Dioxide			6.22E+01	1.51E+03
Furfural			4.59E+00	6.58E-01
Hydrazine			3.89E+00	2.12E+00
Xylitol				
Methane			8.60E+00	
Nitrogen Dioxide			4.60E-02	
Carbon Monoxide			2.80E-03	
Sulphuric Acid			1.75E-01	
Water Effluent	310256.4	kg/hr		

Table A 14: GFT Process

Air Emission Data				
Constituent	Emission Rate (kg/hr)			
	Distillation	Refinery	Boiler	Cooling Tower
Carbon Dioxide	9.11E+02	1.10E+03	7.72E+04	
Oxygen			2.43E+04	
Nitrogen			2.99E+05	
Steam	7.46E+02	8.96E+02	1.94E+04	1.05E+02
Sulphur Dioxide				
Nitrogen Oxides	6.35E-01	9.09E-02	5.38E+01	
Methane				
Carbon Monoxide	1.63E-01	2.12E-01	3.30E+02	
Particulates	3.34E-06	1.80E-01		
Hydro-Carbons		3.01E-03		2.56E-02
Aldehydes		2.58E-04		
Aromatics				1.86E-01
Phenol				5.29E-05
Hydrogen Chloride				6.06E-02
Hydrogen			3.11E+00	
Water Effluent	133055	kg/hr		