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Biomass to Liquids

1 Executive Summary

The purpose of this study was to develop cost curves (excluding feedstock costs) for the conversion of liquid fuels from biomass through the gasification to Fischer-Tropsch pathway on the basis of a 2000-20,000 tonne/day input of biomass. This has been achieved through a techno-economic analysis deriving high level mass and energy balances for the process and order of magnitude estimates of capital and operating costs.

The biomass gasification to Fischer-Tropsch plant is comprised, at the high level, of the following operations:

- Biomass Preparation,
- Gasification,
- Gas Cleanup,
- Fischer-Tropsch Synthesis, and
- Product Upgrading.

In addition to these operations the plant also has an air separation unit to provide oxygen for gasification and nitrogen for biomass preparation; a water gas shift reactor to tailor the syngas composition; and a heat management system to maximise the thermal efficiency of the plant.

The base case plant size was taken as 10,000 oven-dry tonne/day biomass input, with this plant size producing approximately 10,500 barrel/day of Fischer-Tropsch fuel. This is the equivalent of 577 barrels of petrol, 1,497 barrels of kerosene and 8,424 barrels of diesel. The output per oven dry tonne is 0.06 barrels of petrol, 0.15 barrels of kerosene and 0.8 barrels of diesel.

The baseline production cost curve for the plant, excluding the biomass feedstock cost, is shown in Figure 1. The equation for this cost curve is:

$$\text{cost} = 40 + \frac{10000}{p^{0.852}}$$

Where the cost is expressed in New Zealand Dollars and p is the input in tonne/day.

The cost curve has been found to be sensitive to the cost of electricity, the capital cost and the operating cost.

This study has shown that the production cost, excluding feedstock cost, of liquid fuels derived from biomass through the gasification to Fischer-Tropsch pathway is similar to those derived from lignite. Further analysis of the cost and availability of biomass feedstock is required.

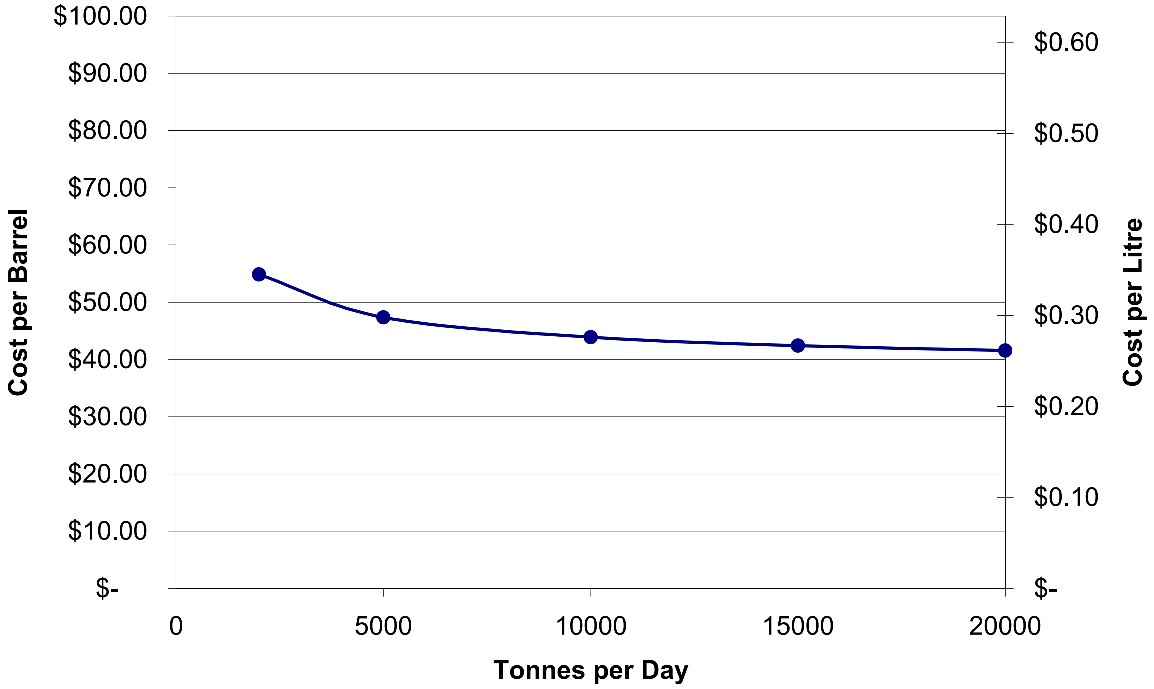

Figure 1: Production Cost Curve for making Liquid Fuels from Biomass via the Gasification to Fischer-Tropsch Pathway. The Plant Size is defined in Terms of Oven-dry Biomass Input and the Cost is the Cost of the Upgraded Fischer-Tropsch Product.

Biomass to Liquids

Contents

2 Introduction

The production of transportation fuels in New Zealand and abroad is based primarily on crude oil. Crude oil is a global commodity with very low price elasticity. This means that prices can fluctuate substantially as a result of small differences between supply and demand. Crude oil is also a finite, non-renewable, energy resource with the exact world supply unknown. The supply of crude oil is also not distributed evenly around the globe meaning that some regions are very rich in crude oil while some are poor. These factors mean that economies, particularly those with little or no oil resources, are motivated to find alternative options for production of transport fuels that offer:

- Stable pricing,
- Long-term security of supply,
- Energy self-sufficiency.

The generation of fossil carbon dioxide from the production and use of transport fuels derived from crude oil also presents a problem to economies where the potential environmental cost of such emissions is converted into an economic cost. One way of overcoming this particular problem is through the development of renewable fuels derived from biomass. Biomass resources are also widespread and therefore potentially offer energy self-sufficiency and security of supply to many economies.

Reasons for considering biomass to liquids as an option for New Zealand include:

- Good growing conditions with a long history in forestry and agriculture,
- Large land area available that is suitable for biomass production,
- Potential synergies with abundant lignite resources to allow a transition from fossil fuels to biomass fuels.

The potential barriers to the uptake of biomass to liquids in New Zealand include:

- Insufficient feedstock for a commercial scale plant,
- The technology is not proven on the commercial scale.

There are a number of methods for producing liquid fuels from biomass feedstocks. One of these methods, the combination of gasification and Fischer-Tropsch is a thermo-chemical pathway for the production of liquid fuels from a variety of solid carbonaceous feedstocks. In this pathway the carbonaceous feedstock is first broken down by gasification into a syngas made up of the basic chemical building blocks of carbon dioxide, carbon monoxide and hydrogen. These building blocks are then reassembled into organic molecules via the Fischer-Tropsch process to produce liquid fuels. This pathway has been studied extensively as an option for the production of transport fuels from coal and lignite including in the New Zealand context (Garrood and Clemens, 2007). This is also considered to be a potential pathway for the conversion of biomass into liquid fuels.

2.1 Fischer Tropsch Fuels

The properties of Fischer-Tropsch (FT) fuels derived from biomass are the same as those derived from coal or lignite as the properties of the feedstock are lost when it is converted into syngas. Garrood and Clemens (2007) considered the properties of FT fuels compared to their petroleum derived equivalents in detail. Their discussion is reproduced here.

"The synthetic petrol and diesel produced by FT processes require no further refining and are direct replacements for conventional petrol and diesel fuels; no modification is necessary to distribution infrastructure or engines and they can be blended with standard petrol or diesel to any ratio. Table 1 shows some properties of FT diesel and petrol compared to Number 2 diesel.

Table 1: Physical properties of FT fuels compared to Number 2 diesel [NREL 2003]*.

	Number 2 diesel	Low temperature FT (FT diesel)	High temperature FT (FT petrol)
Energy content (HHV), MJ/kg	43-48	45-48	45-48
Density at 15°C, kg/l	0.82-0.85	0.77-0.79	0.80
Cetane number	44.9	>74	~50
Sulphur content, ppm	300	<1	<1
Total Aromatics, %	~30	0.1-2.0	~10
Hydrogen wt%	13-13.5	~15	~14.4

*Current New Zealand requirements are 50 ppm sulphur (max) and minimum cetane number 47

The energy density per kilogram of FT fuel is similar to petrodiesel but its lower volumetric density results in an overall lower energy density per litre for FT fuels. The difference is at most 10% and may be as low as a few percent for some individual fuels. Where it is possible to do so, a relatively minor adjustment to the engine management and fuel injection systems should allow similar performance to be achieved. From the average values of energy content and density, it was calculated that the average energy density of the FT fuels was 36.5MJ/litre (equivalent to 27.4litre/GJ), compared to around 38.3 MJ/litre for petrodiesel.

The cetane number of FT diesel is much higher than for standard diesel; above a cetane number of 55, diesel engine noise and NOx emissions are significantly reduced using modern engines [EPA 2002]. In practical studies the NOx emissions were reduced by 13% on average by using FT diesel [NREL 2003].

The fuel may have low lubricity, which can be rectified by the addition of fuel additives as used with low sulphur fuels from crude oil.

The concentrations of aromatic compounds, sulphur and high chain length hydrocarbons in FT fuel are much lower than for standard petrodiesel. As a result lower exhaust particulate emissions are expected (to the order of 26% lower in tests reported [NREL 2003, EPA 2002]).

Carbon monoxide emissions are also observed to decrease with FT diesel, but the results for hydrocarbon emissions have been found to be more variable [NREL 2003].[1]

Because of the favourable attributes of FT diesel, it can be considered as a premium fuel and may have greater value in countries with particular focus on emissions from vehicles."

2.2 Scope of Study

The biomass to liquids plant in this report is based on the lignite to liquids plant in Garrood and Clemens (2007) with the following modifications:

- The system boundary begins at the gate so the feedstock costs are not considered in the analysis,
- The working range of the plant is 2000-20,000 oven-dry tonnes of biomass per day input,
- Un-reacted FT output is recycled.

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[1] Care is required in interpreting these results as responses are very much engine and test dependent and conclusions should be drawn from a much wider sample of tests. Unfortunately there does not appear to be significant data available from which to do this.

3 Methodology

This study has been carried out in three broad segments of work:

- Literature survey,
- Techno-economic analysis, and
- Interpretation and commentary.

The literature survey was a brief analysis of recent studies on biomass gasification and the biomass- FT pathway to ensure the most relevant process configuration was used and most current information was used for determining the mass and energy balances.

The techno-economic analysis comprised of the design of a block flow diagram for the process, identifying key process and process streams. This block flow diagram was then used to determine high level mass, energy and utility balances. These balances in turn provided sizing information for the main capital items in the process providing a basis for factorial capital cost estimation for the process. The sensitivity of the economic analysis to key variables was also determined.

The interpretation and commentary considers the likely impact of key assumptions and sensitivities in determining the overall viability of biomass to liquids via Fischer-Tropsch in the New Zealand context considering alternative options.

4 Literature Survey

Biomass to liquids via FT is a pathway that has been considered in the International literature though no detailed analyses of the process in the New Zealand context were located. The process uses similar unit operations to the coal (or lignite) to liquids pathway though there are differences in the fuel preparation, gasification and cleanup operations.

The main gasification technologies proposed for the biomass to liquids pathway are fixed bed (FB), circulating fluidised bed (CFB) and entrained flow (EF) gasification. The latter is the primary technology currently considered for coal to liquids pathways and for this reason has been chosen as the technology used in this analysis to allow the most direct comparison with the Garrood and Clemens study. Fluidised bed gasification is traditionally the preferred choice for biomass gasification but for the purposes of this study would not allow an effective comparison with the Garrood and Clemens study.

EF gasification has been used in International studies on the biomass to liquid fuels pathway, including van Vliet et al (2009), Uslu et al (2008) and Bergmann et al (2005). The key issue found with using biomass in EF gasifiers is the tenacity of the feedstock in the size reduction processes prior to gasification. EF gasification requires the feedstock to be reduced to approximately 200 µm before injection into the gasifier. A number of pre-treatment steps have been investigated to reduce the energy requirements for preparing biomass for EF gasification, including liquefaction, pyrolysis and torrefaction (Svoboda et al, 2009). The work by both Uslu et al (2008) and Svoboda et al (2009) suggests that torrefaction is the best pre-treatment option.

Torrefaction is a roasting process that substantially breaks down the hemicelluloses in the biomass to cause structural weakening. This has been found to make biomass gasification more efficient by greatly reducing the milling power consumption (by 70-90%) while retaining a substantial amount of the biomass energy in the torrefied wood product (Bergman, 2005; Prins et al, 2006). CRL Energy has undertaken a techno-economic analysis of the Torrefaction process in the New Zealand context as part of its gasification research programme (McCurdy and Williamson, 2009).

5 Techno-economic Analysis

5.1 Process

The key unit operations in the biomass to liquid fuels via FT are:

- Biomass preparation by torrefaction and milling,
- Torrefied biomass gasification in EF gasifier,
- Syngas cleanup,
- Syngas balancing by water gas shift,
- Conversion of syngas to hydrocarbons by Fischer-Tropsch,
- Upgrading of hydrocarbons to fuels,
- Air separation to provide oxygen for the gasifier and nitrogen for torrefaction.

These unit operations are represented as a block flow diagram in Figure 2 with an additional heat management system included to take energy from exothermic processes, such as gasification and FT, and provide it to endothermic processes, such as biomass preparation and hydrocarbon upgrading.

5.1.1 Description of Unit Operations

The following is a detailed description of the unit operations and their operating parameters.

Biomass Preparation

It is assumed that the biomass arrives at the gate in a chipped form as this allows for the easiest transportation and handling and also eliminates some of the differences between potential biomass feedstocks. Where specific biomass properties are required it has been assumed that the biomass is chipped plantation *Pinus radiata*.

Once onsite the biomass immediately enters the torrefaction process. This process actually comprises a drying step followed by a torrefaction step, with the drying step carried out in an air atmosphere at approximately 100°C and the torrefaction step carried out under nitrogen at 280-300°C. The synergy of combining the torrefaction process with EF gasification is apparent with the use of nitrogen from the air separation unit.

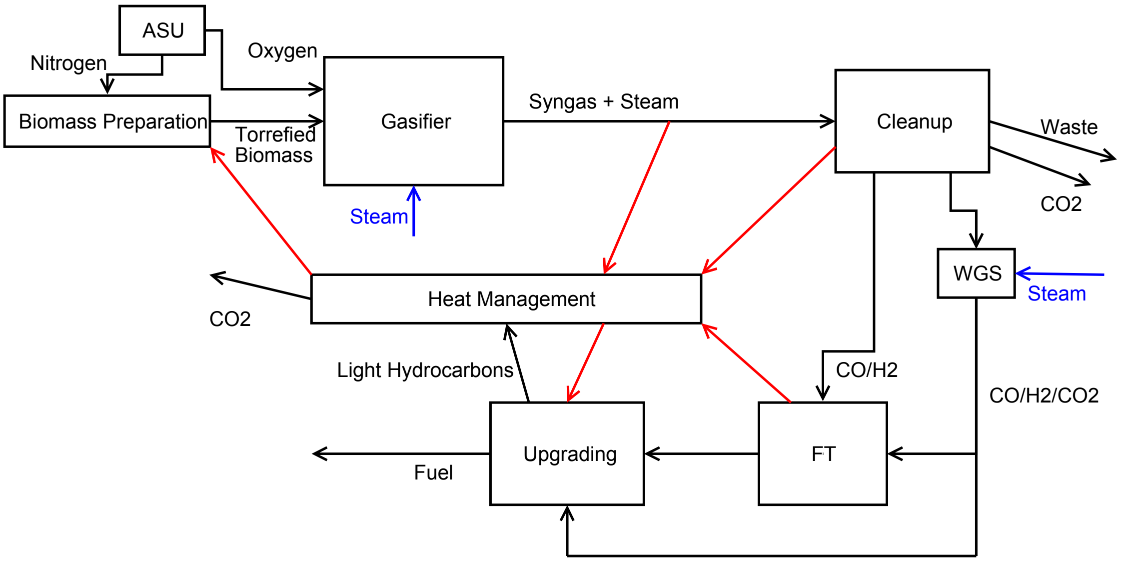

Figure 2: Block Flow Diagram for the Biomass to Liquids via EF Gasification and Fischer-Tropsch process

Operating Parameters

The operating parameters for the biomass preparation operation have been derived directly from McCurdy and Williamson (2009) and are summarised in Table 2. The details from this study provided in Appendix B also show a cooling and natural gas requirement. Both of these are eliminated through the integration with the gasification and FT processes. The nitrogen is also effectively free as a by-product from the ASU.

Table 2: Operating Parameters for Biomass Preparation

Parameter	Value
Mass Efficiency	79% (dry basis)
Electricity	48kW/oven-dry tonne/h
Nitrogen	0.1t/oven-dry tonne/h
Heating	500kW/oven-dry tonne/h

Economic Parameters

The capital expenditure has also been derived directly from McCurdy and Williamson (2009) and is given by the following equation:

$$CAPEX = 340{,}000 \times input + 462{,}000$$

Wherein the CAPEX is in NZ$ and *input* is the annual input in oven-dry tonnes.

Entrained Flow Gasification

The EF gasifier is based on the Shell gasifier. This is a slagging gasifier in which the feedstock is injected as small particles mixed with a pressurised stream of oxygen into the gasifier operating at 1300°C. At this temperature the ash in the biomass forms a slag that runs down the walls of the gasifier. The result is almost complete conversion of the feedstock into syngas, which exits out the top of the gasifier.

Operating Parameters

The operating parameters of the EF gasifier, derived from van Vliet et al (2009) are shown in Table 3.

Table 3: Operating Parameters for the Entrained Flow Gasifier

Operating Parameter	Value
Temperature	1300°C
Pressure	40 bar
Cold gas efficiency	75%
Steam injection	0.2kg/kg of biomass input
Output H_2:CO ratio	2

Economic Parameters

The minimum scale specified for the process is approximately equivalent to the maximum size for an EF gasifier. This means that the scaling of the gasifier operation is effectively linear over the range in question. The CAPEX cost used was $120 million for a 2000 tonnes/day gasifier scaling linearly.

Syngas Cleanup

Syngas is cleaned in a series of processes including:

1) Cyclone
2) Heat exchanger
3) Oil based cleanup
4) Scrubber NaOH
5) Scrubber H_2SO_4
6) Guard beds
7) CO_2 Removal with Selexol

High pressure cleanup is used to reduce the need for compressors and to match the gasifier and FT pressures. The cyclone removes ash and other solids while the oil Scrubber removes tar. To avoid poisoning the catalysts in the FT and water gas shift reactors, a comprehensive scrubbing system follows including units to scrub HCN, NH_3, H_2S and COS. This takes place at a lower temperature (around 100°C).

In order to increase the efficiency of the FT reactor, a CO_2 scrubbing system is included. This is economically viable at higher biomass costs but becomes less so at lower costs. The CO_2 scrubbing market is dominated by the Selexol process which removes a large proportion of the CO_2 from the gas stream. Following this the gas is recompressed in preparation for the FT reactor.

Operating Parameters

Temperature: 1400-100°C

Pressure: 20 bar

Heat exchanger capacity: 437 MW for the 10000 tonne per day

Gas compressor size: 3.7 MW for the 10000 tonne per day

Economic Parameters

Hamelinck (2004) records a series of costs totalling 325 million dollars for a 10,000 tonnes per day plant. This is composed of \$42 million for the general cleanup scaling with a power factor of 0.7, \$16 million for the heat exchanger scaling with a power factor of 0.6 and \$220 million for the CO_2 removal scaling with a power factor of 0.7.

Fischer-Tropsch Reactor

The FT process uses a catalyst to reform $H_2 + CO$ to $(CH_2)_nH_2$ chains + CO_2. A value called the alpha constant characterises the chain growth probability and this changes with temperature and catalyst type. A higher alpha constant leads to longer chain lengths and hence a higher proportion of the heavier hydrocarbons (diesel and waxes). The approximate distribution can be calculated from the alpha constant using the equation:

$$w_i = \alpha^{i-1}(1-a)^2 \times i$$

where w_i is the probability of getting a chain of length i

A slurry phase FT reactor operating at 60 bar and 180-200°C was chosen for this study to achieve the highest realistic alpha constant. The output was calculated using the "Anderson-Schultz-Flory" (ASF) distribution calculation (www.zero.no) with an alpha of 0.92. A major simplifying assumption is made in that all the chains are treated as alkanes. There is an assumption that 10% of carbons are converted into CO_2. With these parameters a heat production rate of 223 MW is derived for the 10,000 tonne/day plant at a temperature of around 180°C.

Operating Parameters

Pressure: 60 bar

Temperature: 200°C

Using an alpha value of 0.92 a chain length distribution with a peak at 12 is obtained as shown in Figure 3.

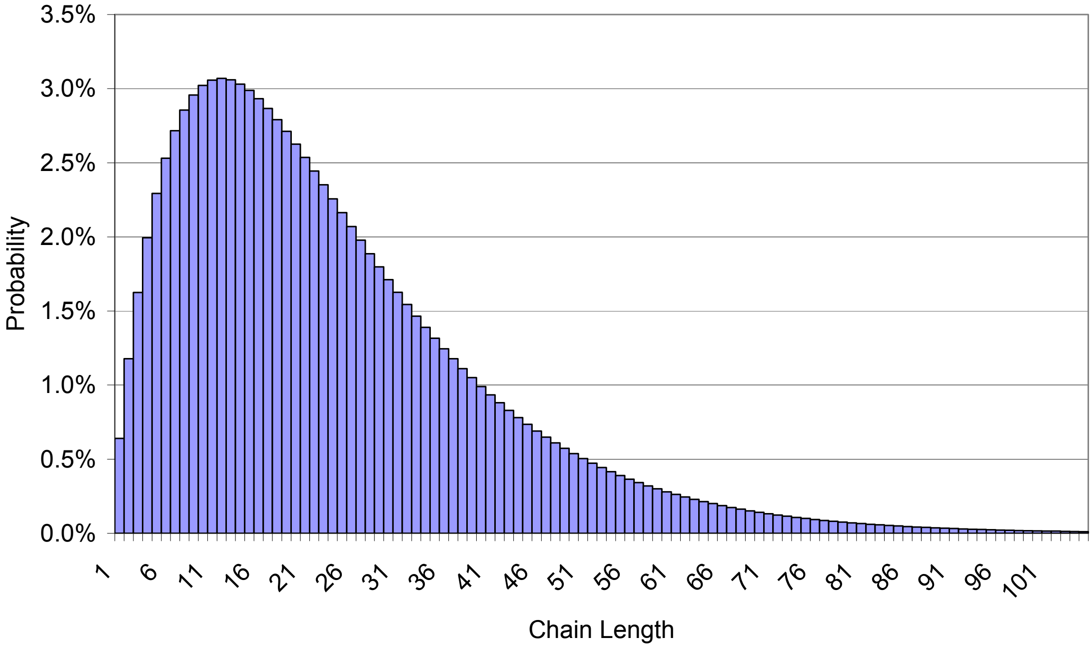

Figure 3: Chain Length Distribution from the Fischer-Tropsch Reactor

Economic Parameters

Costs are based on Hamelinck (2004) with a base of $79 million at 10,000 tonnes per day input and a scaling factor of 0.72.

Fuel Upgrading

The range of hydrocarbons produced from the FT reaction is widely spread, and is now upgraded to contain the more useful parts. The final output emphasises the Diesel part of the spectrum of hydrocarbons. A simplified model of cracking was used where all wax chains were halved until no more remained. Isomerisation, fractional distillation and hydro-cracking combine to create a high quality diesel fuel as well as a variety of kerosene and petrol length products which, although low in octane number provide a useful starting product for a variety of end uses.

Operating Parameters

The operating parameters of this process are variable but were not required for the level of analysis in this study.

Economic Parameters

The capital cost is $32 million at 10,000 barrels per day scaling with a power factor of 0.7.

Air Separation Unit

The air separation unit achieves an oxygen purity of 99% and supplies streams of both oxygen for the gasifier and nitrogen for the torrefaction process. The energy needed to generate one tonne of oxygen is 160 kilowatt hours (576 kJ per kilogram) (Tranier *et al.*, 2009)).

Operating Parameters

Energy: 576 kJ per kilogram

Economic Parameters

A capital cost of the air separation unit (in $US) is given by:

$$\text{cost} = 11248 \times \textit{production} + 2{,}000{,}000$$

where *production* is the oxygen production in tonne/day.

Heat Management System

The heat management system is a catchall for the various items of plant that will be used to integrate the heat flows in the plant for maximum thermal efficiency. The cost of the heat exchange components, for heating or cooling, are assumed to be contained within the auxiliary plant factors used in capital cost estimation for the main operations that produce or use large amounts of heat. The production of heat in the exothermic operations is sufficiently greater and at higher temperatures than the heat that is consumed by the endothermic operations to allow the assumption that all of the process heating requirements can be provided from waste heat. Through a simple recycling system the minimal additional cooling required is supplied from the water supply. The cost of the heat management system integration is included in the cleanup step.

5.2 Process Modelling

The process was modelled in a custom excel spreadsheet using simple feedback loops and gross conversion numbers as detailed above for each of the main processes. A feed forward model simplified the calculations to avoid the need for iterative solving techniques.

5.2.1 Mass Balance

The mass balance for the process is presented in Table 4 on the basis of a 10,000 oven-dry tonne per day plant. This table only shows the major streams with the flows in kg/s and the components expressed in mass fractions. Assuming moisture content of 1kg/kg (50%) the green biomass input is 20,000 tonne per day.

The output per oven dry tonne is 0.06 barrels of petrol, 0.15 barrels of kerosene and 0.8 barrels of diesel. The 84% reduction in mass results from the compression of energy per volume and weight advantage achieved via liquid hydrocarbon fuels along with the energy losses through the process.

Water Demand

The water demand for the process is expected to be minimal as the FT process produces more water than is required by the gasification and water gas shift processes combined. Cooling water is also not a significant utility in this analysis and therefore unlikely to contribute much to water demand. .

Table 4: Mass Balance for the Production of Liquid Fuel from Biomass via Gasification and Fischer-Tropsch

	Flow at 10000 tonne/day input (kg/s)	**Mass Fraction**														
		Biomass	Torrefied Biomass	O_2	H_2O	H_2	CO_2	CO	C1-C4	C5-C7	C8-C10	C11-C12	C13-C20	C21-C30	C31-C60	C61+
Input	116	100%														
Torrefied Biomass	91		100%													
Oxygen input	92			100%												
Gasifier input	201		45%	46%	9%											
Gasifier output	183				7%	3%	68%	22%								
Cleanup output	45					13%		88%								
Bypass	43					13%		88%								
WGS In	6				63%	5%		33%								
WGS Out	6				46%	7%	41%	7%								
Fischer-Tropsch In	46					12%	5%	82%								
Fischer-Tropsch Out	46				47%	1%	13%		0%	1%	1%	1%	6%	9%	16%	4%
Upgrade In	18					1%			1%	2%	3%	3%	16%	22%	41%	11%
Upgrade Out	18								1%	2%	3%	14%	71%	9%		

5.2.2 Energy Balance

The main energy flows in the process are shown in Table 5. The only significant energy inputs are the biomass and electricity. All of the heat requirements of the process are generated from the biomass input. The overall efficiency of the process is 42% meaning that 42% of the energy in the biomass and electricity inputs is contained in the final product. The additional energy gained in the product upgrade operation comes from heat input.

Table 5: Process Energy Flows

	Energy of flow (MW)
Biomass input	1736
Total electricity input	77
Total Energy input	**1813**
Torrefied Biomass	1638
Gasifier output	1201
Fischer-Tropsch Out	666
Upgrade Out	**780**
Total Efficiency	43%

5.3 Economic Analysis

Table 6 shows the general economic parameters used throughout this study.

Table 6: Economic Parameters

Parameter	**Value**
Interest Rate	10%
Loan Period	10 years
Electricity Price	6 c/kWh
NZD:USD	1.30
NZD:EUR	1.98
Availability	8400 hours/year

5.3.1 Capital and Operating Costs

The breakdown of capital cost estimates for a 10,000 oven-dry tonne/day plant is shown in Table 7. The total cost for the plant is NZ$1.49 billion with the total capital cost for the plant sizes in the study ranging from $336 million for a 2000 tonne per day plant to $2.87 billion for a 20,000 tonne/day plant.

Table 7: Capital Cost Estimate for a 10,000 tonne/day Plant

Unit Operation	Capital Cost Estimate
Biomass Preparation	$555 million
Air Separation Unit	$119 million
Gasification	$472 million
Cleanup	$223 million
Water Gas Shift	$9 million
Fischer-Tropsch	$79 million
Product Upgrading	$32 million
Total	$1,488 million

The operating costs for the biomass to liquids process for different plant sizes are shown in Figure 4. The relationship is approximately linear. Table 8 shows the operating expense breakdown for the base case.

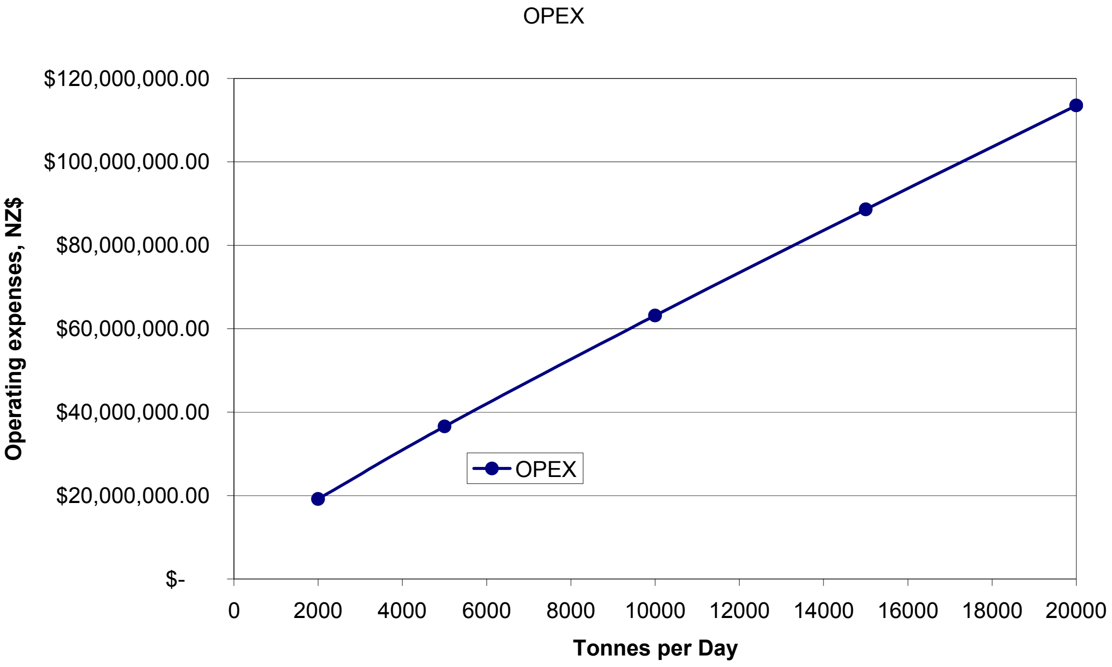

Figure 4: Operating Expenses for Biomass to Liquids Plant

Table 8: Opex Breakdown

Opex (NZ$)	
Maintenance	$44 million
Labour	$11 million
Gas Cleaning (wet)	$7 million
Electricity	$38 million

5.4 Cost Curve

The cost curve for the production of liquid fuels from biomass using the gasification to FT pathway is shown in Figure 5. The equation for the line fitting this curve is:

$$\text{cost} = 40 + \frac{10000}{p^{0.852}}$$

Wherein the cost is expressed in New Zealand Dollars and p is the input in tonne/day.

There are 159 litres per barrel so the cost per litre ranges from 36 c/l at 2000 tonne/day to 26c/l at 20,000 tonne/day.

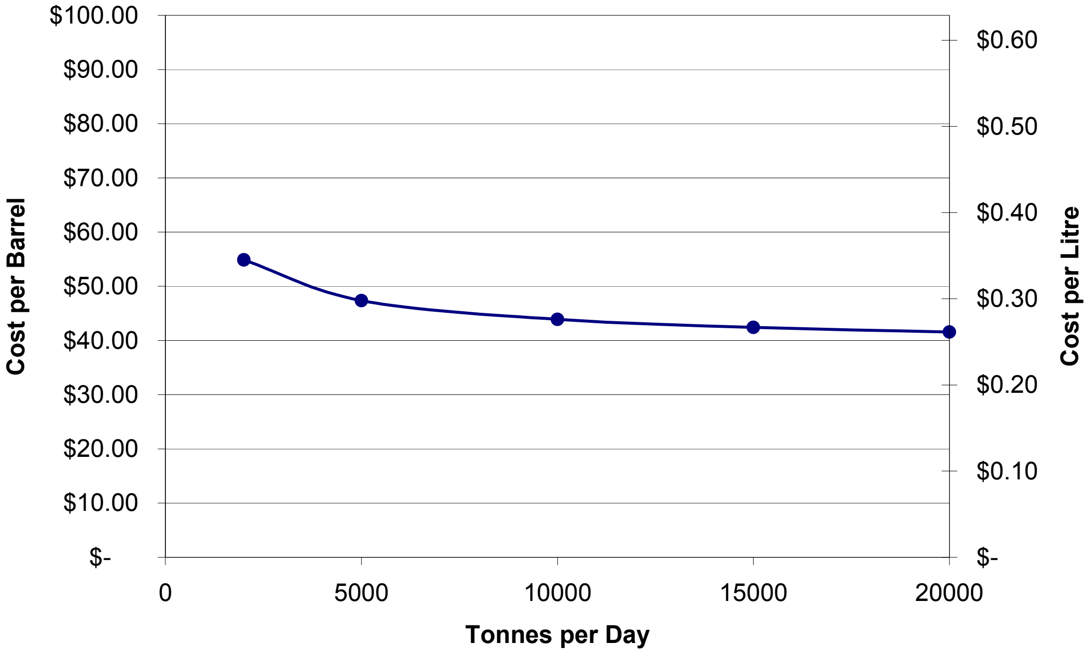

Figure 5: Cost of Production of Liquid Fuels from Biomass via Gasification and Fischer-Tropsch

5.4.1 Cost Curve Sensitivities

The cost curve is sensitive to variations in the electricity cost, the CAPEX cost and the OPEX cost. These sensitivities are expressed in Figure 6 to Figure 8. Variation in these parameters shifts the cost curve up or down the ordinate axis.

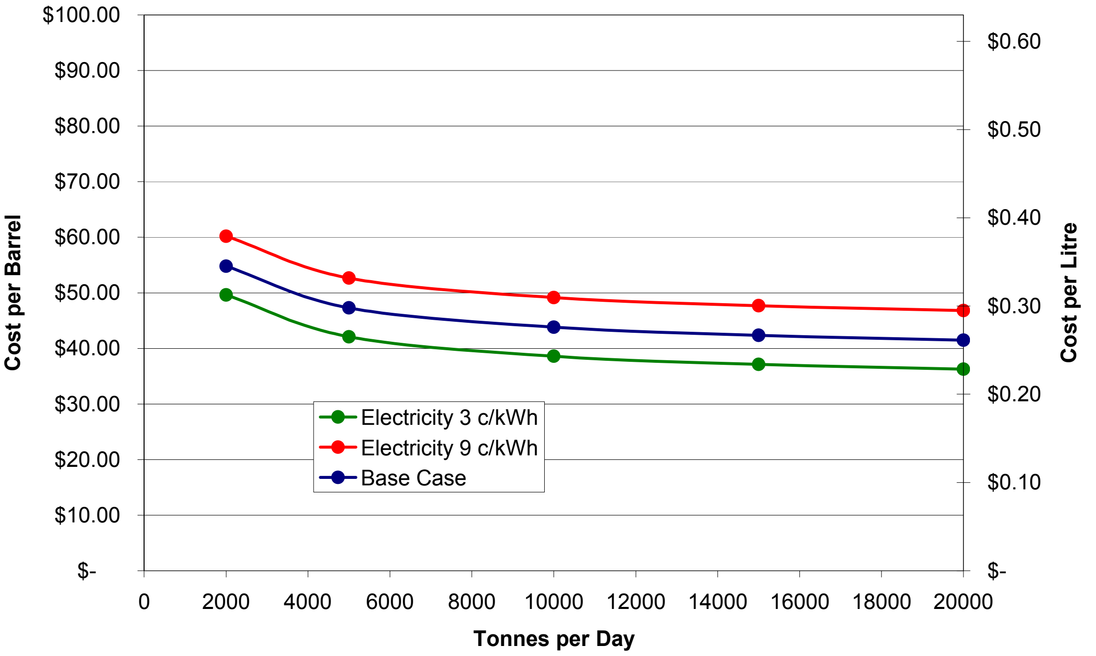

Figure 6: Cost Curve Sensitivity to the Price of Electricity

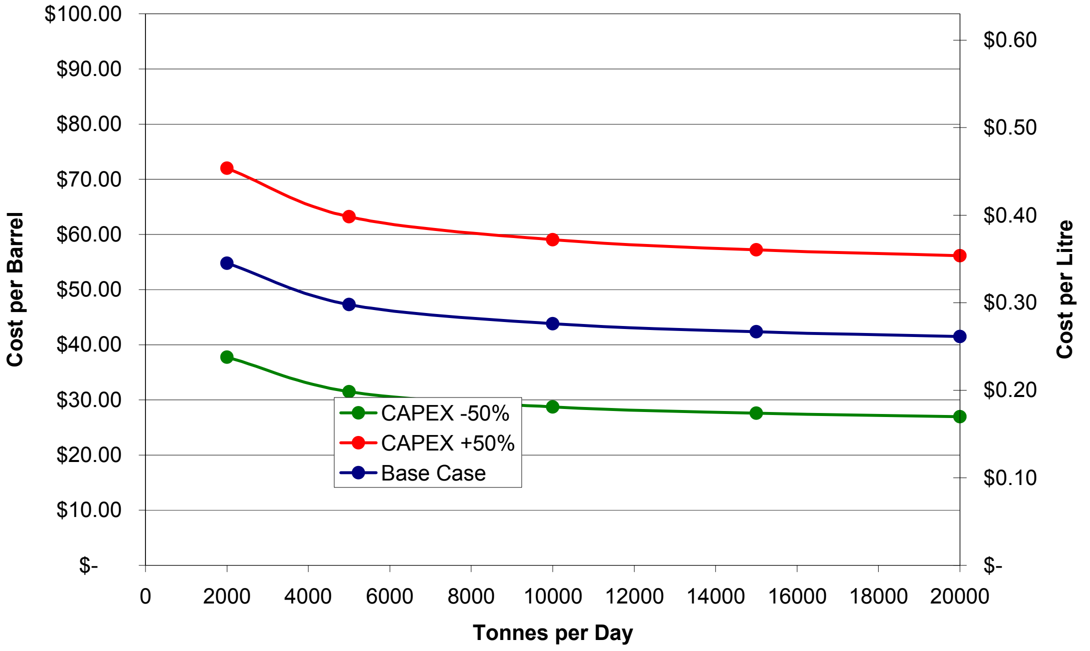

Figure 7: Cost Curve Sensitivity to CAPEX Cost Variation

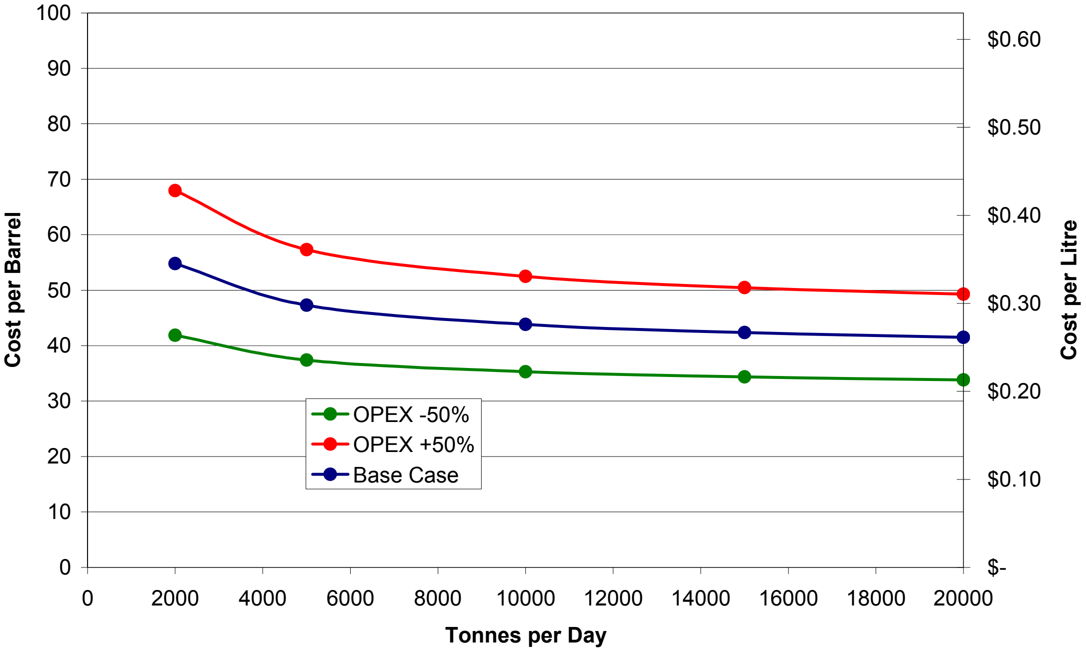

Figure 8: Cost Curve Sensitivity to OPEX Cost Variation

6 Interpretation and Commentary

Garrood and Clemens' study of the production of liquid fuels from lignite found that the cost per litre of fuel produced was 47c/l to 63c/l. The base production cost from biomass determined in this study, of 26c/l to 35c/l, compares favourably to this. The production cost from Garrood and Clemens included the cost of the lignite of approximately 11c/l and if this is removed then the cost of production is 36c/l to 52c/l. On this basis the production of liquid fuels from biomass appears to be cheaper than from lignite but a range of other factors need to be taken into consideration.

Some of the differences between the two sets of results can probably be attributed to the different methodologies used in the two studies as well as changes over time. The Garrood and Clemens study was based on a currency conversion from a US study reported in mid-2006. The present study relies on the most recent data available and will therefore have benefitted from cost reductions and improvements in technology over the last three years.

The Garrood and Clemens study also included CO_2 capture equipment and power block equipment, neither of which was included in this study. The former is specifically not required for biomass while the latter was not considered as an option. Together these two components contribute about 20% of the capital cost of the lignite plant with the power block being the greatest contributor. The gas cleanup for a lignite fuelled plant may be more expensive due to the need to remove pollutants such as sulphur and heavy metals that are usually negligible in ligno-cellulosic biomass.

Based on the discussion so far it is reasonable to conclude that the cost of liquid fuels production from biomass is at least similar to the cost for production from lignite when the feedstock cost is excluded. There is also potential that the cost is slightly lower for production from biomass.

The product from this process is almost the equivalent of refined petrol, diesel and aviation fuel and it should at the very least be considered a high value refinery blend stock rather than a crude. This means that the cost of the product can also be compared to the cost of these transport fuels. Comparison with petrol for example requires the addition of 53c per litre of excise tax, bringing the total cost to 88c per litre. This is below the current cost of 95 octane petrol so there is room for feedstock costs and profit margins to be added. If savings can be made to production costs and feedstock costs kept to a minimum then it may be possible to match the cost of fossil derived transport fuels.

The sensitivity analysis has shown that the production cost is sensitive to the price of electricity, the initial capital cost and the operating costs. The capital cost estimate developed in this study is an order of magnitude estimate and therefore has accuracy to ±50%. There is clearly scope for savings here, though conversely the capital cost could end up being in the upper range. One key area for savings is the complete integration of the biomass preparation operation with the gasification and FT operations, which could reduce the capital cost associated with the preparation operation (currently the largest contributor). Heat integration is likely to be the key to reducing capital cost of the plant.

It is possible for a biomass to liquid fuels plant to generate electricity from waste heat (steam cycle generation) and waste gases (gas turbine). This reduces the overall efficiency of production but will greatly reduce the cost of electricity (effectively reducing it to around 3c/kWh). If electricity generation was to be considered then a detailed analysis would be required as there is potential for the loss of overall efficiency to outweigh the savings from electricity generation.

The scope of this study did not allow for detailed analysis of the operating expenses, so potential for savings in this area are speculative. Overall potential savings from electricity generation and CAPEX savings from process synergies will amount to 5-6c/litre each.

When comparing this study with Garrood and Clemens it should be noted that the cost of carbon dioxide emissions has not been considered in either. Inclusion of this cost will most likely favour the use of biomass over the use of lignite. The capital and operating costs are similar for the two studies. The key difference is the feedstock cost.

6.1 Critical Assumptions

The key critical assumption made in this study was the choice of entrained flow gasification technology over fluidised bed or circulating fluidised bed. Fluidised bed technologies (fixed and circulating) are more tolerant of feedstock variation and can handle a greater feedstock size range than entrained flow. They therefore offer the potential to reduce feedstock preparation costs. Their lower temperature operation however means that they produce tars and aromatic compounds that require more expensive clean up operations and have a lower conversion of biomass into syngas. The fluidised bed technologies also often operate at atmospheric pressure using air rather than oxygen which limits their value for FT synthesis. An unknown area that could prove detrimental in entrained flow gasification is the formation of corrosive slag from the biomass feedstock.

Taking into consideration the overall contribution of the capital cost to the production cost and the contribution of the gasification and cleanup operations to the capital cost, it is unlikely that the choice of gasification technology has had a significant impact on the results of this study.

Related to the assumption of entrained flow gasification is the assumption that torrefaction is used as a feedstock preparation step. There are alternative processes, particularly pyrolysis, that could be used in place of torrefaction but these have been found to result in greater loss of energy.

This study has focussed particularly on the production of FT diesel as the goal fuel rather than petrol. This assumption was made in order to maximise the yield of useful liquid products. If the FT process targeted the petrol fractions then there would be a much higher quantity of lighter hydrocarbons produced. These would either need to be burned as fuel, steam reformed into syngas and recycled into the reactor or recycled back to the gasifier. In all cases the energy efficiency of the process suffers.

Assumptions have also been made on the heat management system to simplify the analysis. The assumption is that there is sufficient heat of sufficient quality generated by exothermic processes to supply all of the heating requirements of endothermic processes. It is possible that this assumption may actually lead to energy wastage.

It has been assumed that the moisture content of the biomass feedstock is 1kg/kg, which is either 50% or 100% depending on the basis. This is probably reasonable for wood chips but may be high for other biomass feedstock. As the basis of the process is oven-dry the only operation that is influenced by feedstock moisture content is preparation. Some variation in the moisture content is unlikely to change the economics of the process significantly as the moisture is removed from the feedstock using waste heat from down stream processes.

7 Conclusions

This study has shown that the production cost of liquid fuels derived from biomass through the gasification to Fischer-Tropsch pathway is similar to those derived from the fossil fuel options, such as crude oil refining and lignite to liquid fuels. More detailed analysis of the feedstock costs is required however to determine the true cost of liquid fuels derived from biomass.

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A.3 Web References

http://www.zero.no/transport/bio/fischer-tropsch-reactor-fed-by-syngas

Appendix B - Torrefaction

B.1 Introduction

The following is an excerpt from McCurdy and Williamson (2009) providing the relevant details for the costing of the torrefaction process.

B.2 Methodology

This study was carried out by determining a base case for the torrefaction process and then applying ranges to key parameters in the base case to analyse sensitivity to variation.

The base case is for a feed into the torrefaction unit of 500,000 tonnes per annum of oven-dried woody biomass. The steps in the base case process are:

- Drying of the wood chips to produce oven-dried wood chips,
- Torrefaction of the oven-dried wood chips to produce torrefied wood,
- Milling of the torrefied wood to produce sub-0.1mm torrefied wood powder.

The drying, torrefaction and milling processes will occur on the same site as the entrained flow gasification plant that will utilise the torrefied wood powder. This allows for integration of the utilities systems for both processes and means that the gases produced during torrefaction can be used for process heating or injected into the gasification process.

B.3 Unit Operations

B.3.1 Drying

The wood chips will be dried using a rotary dryer as this is the standard method for drying wood chips on an industrial scale due to low cost of maintenance and lower specific energy consumption compare to other methods (Meza et al., 2008). Meza et al. (2008) determined the energy consumption for the rotary drying of wood chips and found that 95% of the specific energy utilisation (SEU) was attributed to thermal energy and 5% to electrical energy. The thermal energy directly used to dry the wood chips was 46% of the total thermal energy with flue gas losses the other major contributor as shown in Table 9. The flue gas losses arise in this case because the system tested utilised indirect heating with a natural gas air heater. This loss could therefore be greatly reduced by directly heating the dryer with flue gases. There is no problem with contamination of the product so this is likely to be the best option, especially considering the hot torrefaction gases could be burned to provide the energy. This configuration will roughly halve the energy input to the dryer making the percentage of electrical input increase to 10% and the thermal energy used for drying will increase to approximately 90% of the total thermal energy input.

Table 9: Thermal Energy Balance for Rotary Drying of Wood chips (from Meza et al., 2008)

Item	Percent of Total
Energy Input	100%
Heat for Drying	48%
Flue Gas Losses	47%
Radiation Losses	3%
Leakage Losses	1%
Other Losses	1%

The temperature of the gas inlet to the rotary dryer will be around 200°C as higher temperatures run the risk of auto ignition of the wood chips. The exit temperature from the dryer is expected to be 100°C with the product at 50°C and 10% moisture content (dry basis).

B.3.2 Torrefaction

The torrefaction reactor will be a modified dryer operating at 270°C and utilising recycled torrefaction gases as the inert medium. The torrefaction process will remove all remaining moisture from the wood chips and the solids yield will be 65% (dry basis) with 75% of the energy in the wood chips being retained in the torrefied wood. The remainder of the energy (25%) is in the torrefaction gases. The process energy input to the process (to drive the reactions) is unclear, but from Prins (2005) it appears to be 0.5% to 3% of the energy in the wood chips.

Cooling of the torrefied wood is also required before it is exposed to air to prevent auto ignition.

B.3.3 Milling

The torrefied wood will be milled in a high speed hammer mill with the energy requirements of this process being 10kW of electricity per MW of thermal energy contained in the feedstock.

B.4 Heat and Mass Balance

The block flow diagram used for the heat and mass balance for the process is shown in Figure 9. Along with the drying, torrefaction and milling processes described above, this diagram also includes a combustor and a heat exchanger. This takes the gases produced during torrefaction and burns them to provide heat to drive the torrefaction process and the drying process. Additional energy required by the process is provided from the combustion of natural gas in the same combustor. Heat is supplied to the torrefaction unit through the heating of a recycle stream (6 and 7) from the torrefaction reactor. The flow of the recycle stream is 8 times that of the purge stream (8). The drying process is heated directly by passing the hot flue gases (10), diluted with air (15), through the wood chips.

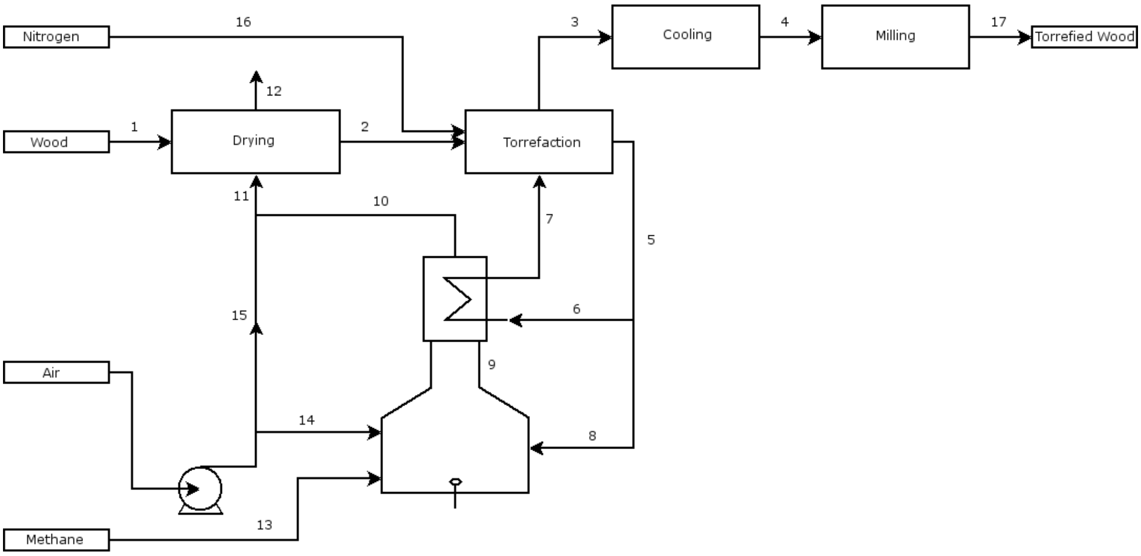

Figure 9: Block Flow Diagram for the Torrefaction Plant

The mass balance for the base case torrefaction process is shown in Table 10. The starting point for the mass balance was stream 2 as the flow of oven-dry wood into the process is set at 500,000 tonnes per annum.

Table 10: Flow and Composition of Streams in the Base Case Torrefaction Process

Stream	Description	Flow, t/h	Composition (mass fraction)									
			ODW	TW	H_2O	CO_2	N_2	O_2	Org	Acetic	CO	CH_4
1	Wood Inlet	119	0.50		0.50							
2	Dried Wood	65	0.91		0.09							
3	Torrified Wood	47		1.00								
4	Cooled Torrefied Wood	47		1.00								
5	Torrefaction gas	142			0.23	0.13	0.19		0.26	0.16	0.03	
6	Recycle	126			0.23	0.13	0.19		0.26	0.16	0.03	
7	Heated Recycle	126			0.23	0.13	0.19		0.26	0.16	0.03	
8	Purge	15			0.23	0.13	0.19		0.26	0.16	0.03	
9	Hot Flue Gas	117			0.08	0.11	0.73	0.07				
10	Flue Gas	117			0.08	0.11	0.73	0.07				
11	Cooled Flue Gas	456			0.02	0.03	0.78	0.18				
12	Dryer Exhaust	510			0.12	0.03	0.69	0.16				
13	Methane	2					0.00	0.00				1.00
14	Combustion Air	109					0.79	0.21				
15	Drying Air	339					0.79	0.21				
16	Nitrogen	3					1.00	0.00				
17	Torrefied Wood Powder	47		1.00								

The energy flows in the process are shown in Table 11 with streams 1 and 13 being energy inlets and stream 17 being the energy outlet. The overall energy efficiency of the process is approximately 83% (845/(913+108)). This table also shows the temperatures of the streams.

Table 11: Energy Flows in the Base Case Torrefaction Process

Stream	**Description**	**Temperature, °C**	**Cp, kJ/kgK**	**Energy Flow, GJ/h**
1	Wood Inlet	20	2.13	913
2	Dried Wood	120	0.80	1034
3	Torrified Wood	270	1.15	849
4	Cooled Torrefied Wood	150	0.68	849
5	Torrefaction gas	270	2.15	869
6	Recycle	270	2.15	774
7	Heated Recycle	350	2.15	774
8	Purge	270	2.15	95
9	Hot Flue Gas	1251	1.51	0
10	Flue Gas	1129	1.51	0
11	Cooled Flue Gas	300	1.51	0
12	Dryer Exhaust	150	1.39	0
13	Methane	20	2.19	108
14	Combustion Air	20	1.01	0
15	Drying Air	20	1.01	0
16	Nitrogen	180	1.04	0
17	Torrefied Wood Powder	150	0.68	845

B.4.1 Utilities Requirements

The utilities requirements for the process are shown in Table 12. The electricity requirements were determined from the characteristics of the process equipment. The cooling requirements were determined from the heat contained in the streams that required cooling. The nitrogen and methane requirements were determined directly from the mass balance.

Table 12: Utilities Requirements for Torrefaction Process per Hour and Daily

Process	**Electricity, kW**	**Cooling, kW**	**Nitrogen, t/h**	**Methane, GJ/h**
Drying	357	0	0	0
Torrefaction	167	0	3	0
Cooling	0	2709	3	0
Milling	2359	2359	0	0
Combustion	0	0	0	107.5
Total	**2883**	**5068**	**6**	**108**
Daily Requirements	**69MWh**	**122MWh**	**143t**	**2580GJ**

B.5 Economic Analysis

B.5.1 Capital Cost Estimation

The capital cost of the torrefaction plant was estimated by determining the cost of the main plant items (MPIC) and then multiplying the sum of these by a Lang factor of 4.3 which is standard for a solid-fluid processing plant. Innovative step factors were not applied as the torrefaction process is assumed to have enough similarities to drying so that no process innovation is required. The costs main plant items along with the capital cost estimate are shown in Table 13.

Table 13: Capital Cost Estimate for the Base Case Torrefaction Process

Plant Item	**Size**	**Cost, NZ$000 (Dec 2008)**	**Notes**
Dryer	$1595m^2$	$9.455	Carbon steel rotary dryer, converted from USD(2007)
Blower	134kW	$171	Centrifugal
Torrefier	$503m^2$	$2,471	Carbon steel rotary dryer, converted from USD(2007)
Burner	61589kW	$4,177	Based on gas-fired steam boiler
Heat Exchanger	$138m^2$	$87	Floating head shell and tube
Mill	2359kW	$198	Based on rod and ball mill
Total		$16,561	
Lang Factor		4.3	For solid-fluid plant
Estimate		**$71,215**	

The capital cost estimate represents a specific investment of $280,723 per MW_{th} input. When convert to millions of Euros to compare with Uslu et al (2008) this comes to €0.13M/MW_{th}. By comparison Uslu et al (2008) found a specific investment of €0.16M/MW_{th} but this was for a smaller plant (40MW rather than 254MW) and so the economies of scale are likely to be worse. The range of the estimate by Uslu et al (2008) (0.13-0.19) falls within the error of the cost estimate for this study (0.06-0.19).

B.5.2 Operating Cost Estimate

The operating cost estimate is made up of the cost of capital (interest), depreciation, utilities, labour and maintenance.

Cost of Capital

The cost of capital for the base case was determined using the parameters shown in Table 14. The interest was determined quarterly. The cost of capital calculated for the base case was \$8.27 million per annum with the interest component in this being \$4.7 million per annum.

Table 14: Parameters for Determining the Cost of Capital for the Base Case

Interest Rate	10%
Duration of Loan	20 years
Number of Payments	80

Depreciation

Depreciation was determined as a 20 year straight line resulting in depreciation cost of \$3.56 million per annum.

Utilities

The utilities costs used for the base case are shown in Table 15. These are based on a plant converting 500,000 tonnes of oven-dry wood per year.

Table 15: Utilities Costs for Base Case Torrefaction Process

Utility	Price	Quantity per Day	Daily Cost	Annual Cost
Electricity	\$100/MWh	69MWh	\$6,920	
Cooling	\$1.60/MWh	122MWh	\$194	
Nitrogen	\$50/t	143t	\$7,143	
Natural Gas (Methane)	\$6.90/GJ	2580GJ	\$17,804	
Total			**\$32,061**	**\$11.2 million**

Labour

The labour costs were determined as shown in Table 16.

Table 16: Determination of Labour Costs for Torrefaction Process

Staff per Shift	5
Hours per Year per Shift	1920
Shifts per Year	4.375
Total staff	22
Average Salary	\$70,000
Labour Cost	\$1,540,000

Maintenance

The maintenance cost was taken to be 3% of the capital cost per year for a total of \$1,780,397 per year in the base case.

Total Operating Costs

The total operating costs for the base case torrefaction plant is \$22.8 million with the detailed breakdown of this cost shown in Table 17.

Table 17: Breakdown of the Operating Costs for Base Case Torrefaction Plant

	Cost
Interest	$4,707,667
Depreciation	$3,560,794
Utilities	$11,221,224
Labour	$1,540,000
Maintenance	$1,780,397
Total	**$22,810,102**

B.6 Discussion

B.6.1 Base Case

The conversion rate of the torrefaction process is a significant factor in determining the economic feasibility. A higher rate of conversion, which is inversely proportional to the temperature, reduces the cost of the torrefied wood product. It is unlikely to be feasible to simply lower the torrefaction temperature to maximise the conversion as this will affect other processes, particularly milling, and has an impact on the storage properties of the torrefied wood.

The conversion data used was derived from experiments on willow and therefore may or may not reflect the torrefaction behaviour of particular species used.

B.6.2 Base Case Recommendations

The economic analysis was high level and therefore has considerable scope for improvement. Potential improvements include more clearly defining the operation of the main plant items to give a better indication of capital cost. This will also allow more detailed sensitivity analysis of the utilities requirements of the process.

B.7 References

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