The production of Fischer Tropsch liquids and power through biomass gasification

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The production of Fischer Tropsch liquids and power through biomass gasification

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Colofon

This project has been carried out for the University of Utrecht and Shell Global Solutions International B.V., Amsterdam. Economic evaluations of the full conversion concepts formed part of the SDE project (Samenwerkingsverband Duurzame Energie), in which the University of Utrecht and Shell both participate.

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Preface

This thesis has been written to conclude the doctoral program of the study Science & Policy of the University of Utrecht. The work has been carried out at the University of Utrecht, with the excellent help of A. Faaij and C. Hamelinck and at Shell Global Solutions International B.V. in Amsterdam, with the excellent help of M. Van Hardeveld. I have enjoyed working with them.

The production of CO_2 neutral liquid transportation fuels can contribute largely to the reduction of global CO_2 emissions. Although biomass gasification and Fischer Tropsch are well known research areas, the two combined has had little attention. For this reason, the co-operation between the University and Shell was important to combine the University's knowledge about biomass with Shell's knowledge about Fischer Tropsch.

Experts from many different companies, universities and research centres have helped with the realisation of this report. The author would like to thank G. Grootveld, M. Huisman, F. Loerakker, Klinkenbijl and J. Rajani from Shell Global Solutions International B.V., Amsterdam, M. de Souza from the University of Utrecht and J. Irving (BCL), R. Knight (IGT), K. Salo (Carbona Oy), S. Mitchell (IEA Coal), R. Hogendoorn (FW), K. Daey-Ouwens (ECN), and M. Paisley (BCL).

I have found biomass and Fischer Tropsch a very interesting subject and I hope the reader will also experience this enthusiasm.

Michiel Tijmensen

Summary

The objectives of this study are to identify the technical feasibility and economics of BIG-FT processes in general, to identify most promising system configurations and to identify key R&D issues essential for the commercialisation of BIG-FT technology.
Little work has been carried out so far in the field of BIG-FT (Biomass Integrated Gasification – Fischer Tropsch), though it has been identified as promising.

Different gasification methods produce a wide range of syngas compositions, e.g. with H_2/CO ratios varying between 0.45 and 2. The FT process uses H_2 and CO to form hydrocarbons of different length. In principle, numerous process configurations for the conversion of biomass to FT liquids are possible, e.g. depending on gasifier types, the FT process considered and gas cleaning process. A scheme of the main process steps to convert biomass to FT liquids (and power) and possible different options is shown in figure S.1.

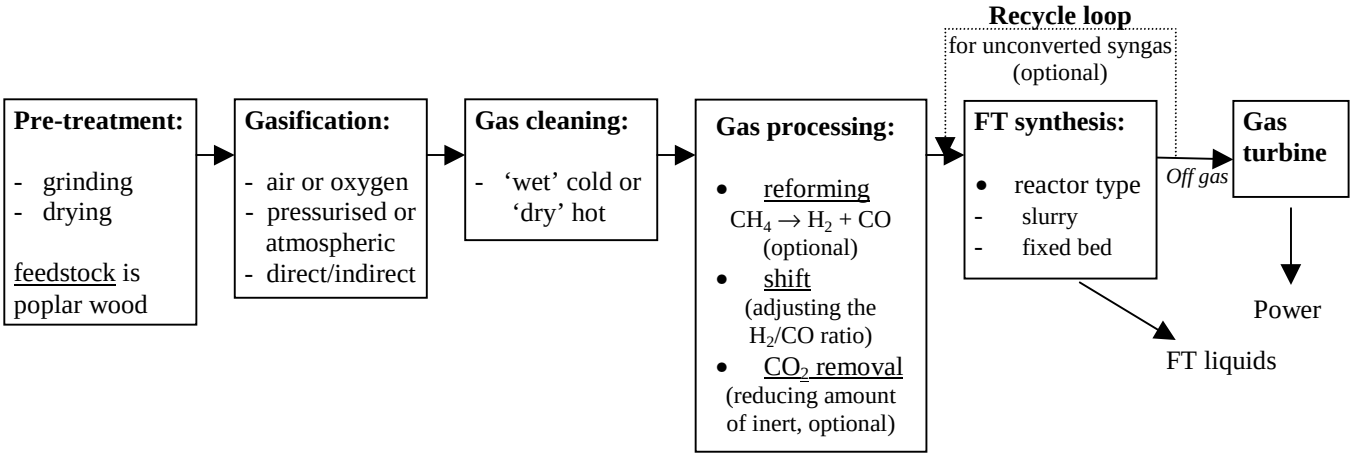

Figure S.1: A schematic view of converting biomass to FT liquids combined with a gas turbine

The work included a technology assessment, system performance calculations with the flowsheet modelling tool Aspen^{plus} and an economic evaluation. To investigate economies of scale, the capacity considered of the BIG-FT systems is in the range of 100 to 1600 MWth.
Various biomass gasification processes (TPS< BCL, IGT, EP) will be studied in combination with FT-concepts in two main categories:

1. Full conversion Fischer Tropsch with the possible use of a gas turbine, focussed on a maximum amount of FT liquids (40% once through conversion).
2. Once through Fischer Tropsch with co-firing of the off gas in a gas turbine (60% and 80% once through conversion).

Pressurised systems have much better *overall energy efficiencies* (42-50% LHV) than atmospheric systems (33-40% LHV). This is mainly due too the high electricity consumption of the syngas compressors when atmospheric gasifiers are used.
Both the IGT and EP gasifier come forward as most suitable for BIG-FT systems. In the FT synthesis high CO conversion, either once through or after recycle, and high C_{5+} selectivity are important for a high overall energy efficiency.

In the *short term*, production costs of FT liquids will be about 35 Hfl/GJ. Investment costs have a share of 50% in overall production costs of FT liquids. The pre-treatment, gasification (with oxygen) and cold gas cleaning account for almost 75% of total capital costs. Biomass costs are 30% of total production costs, O&M (Operation and Maintenance, almost proportionally to investment costs) 20%.

In the *longer term* with large-scale production, high liquid selectivity in FT process, and reduction of investment costs and O&M through technological learning, production costs of FT liquids could drop to below 20 Hfl/GJ. Reduction of capital costs for plants built in the longer term, due too scaling up (12%) and technological learning (15%) have great impact on overall production costs. Biomass costs per GJ FT liquid will decrease due to an increase of overall energy efficiency. Overall energy efficiency will be higher for a third generation plant, due to higher C_{5+} selectivity and higher (once through) CO conversion.

When diesel is the desired final product, the FT product (mainly wax like material) requires *hydrocracking*. In a gasoil mode, the hydrocracking process produces 60% diesel, 40% naphtha and kerosene. Hydrocracking will add about 5% to production costs. FT liquids are totally free of sulphur, nitrogen, vanadium, asphaltenes and aromatics. FT diesel has a very high cetane number and is of excellent quality. FT naphtha has a very low octane number, while FT kerosene in aviation still needs approval of several product specs. Finding a suitable outlet for FT naphtha and kerosene might be a matter of concern.

Conventional *production costs* of diesel are about 0.30 Hfl/liter or 8.4 Hfl/GJ (ex refinery). Production costs of 'green' FT diesel, naphtha and kerosene (35 Hfl/GJ) are not competitive with conventional prices. In the longer term conventional prices could go up due too higher oil prices, but still 'green' FT liquids (19 Hfl/GJ) are not competitive with conventional prices. In the Netherlands, for 'green' electricity a premium is paid between 0.05 and 0.08 Hfl/kWh (on top of the grid price). This corresponds with a premium between 14 and 22 Hfl/GJ. Assuming that a similar premium is paid for 'green' FT liquids, production costs should be between 22 and 30 Hfl/GJ. In the short term pressurised concepts have production costs close to conventional production costs plus premium. In the longer term however, production costs will be lower than conventional production costs plus premium.
In the Netherlands diesel costs are approximately Hfl 1.80 per liter (≈50 Hfl/GJ) at a gas station. Approximately Hfl 0.90 (≈25 Hfl/GJ) per liter is excise duty, received by the Dutch government and another Hfl 0.27 is VAT (www.shell.com, 1999). If 'green', sulphur free FT diesel, naphtha and kerosene would be exempted from excise duty production costs should be below 34 Hfl/GJ. In the short term BIG-FT systems are very close to these production costs, in the long term production costs are likely to be lower.

There are several uncertainties with respect to the *technology status*. A very critical step in the whole system is gas cleaning. It still has to be proven if the gas cleaning section is able to meet the strict cleaning requirements for FT synthesis. Possibly Sulfinol D is required for cleaning purposes, thereby raising production costs. Pressurised (oxygen) gasification systems, having most promising economics and advantages of scale, still need further development. At present, only atmospheric air gasification systems, operating at relatively small scale, have proved to be reliable.

The used *syngas compositions* as produced by the different gasifiers have strong influence on overall results. Most data are taken from literature and based on pilot-scale operating experience. The reliability of these data for large-scale gasifiers is not known.

In the long term the overall energy efficiency of the concepts will be higher if high selectivity can be combined with high conversion. This could be realised in either fixed bed or slurry reactors. Costs for slurry reactors, which are not available yet, could be lower than for fixed bed reactors and will definitely have better economies of scale. Heat integration can also be improved. Power generation in the gas turbine will improve if used on large scale. Hot gas cleaning has high potential of improving efficiency, but the uncertainties about developments and costs of this promising option are substantial.

Recommendations for further actions and research are:

- The gas cleaning section needs special attention, due too the sensitivity to contaminants of the FT catalyst. Proper data sets of contaminants in the syngas must be made, with high detection accuracy. Hot gas cleaning is promising, but will require even more development before sufficient cleaning is guaranteed.
- Pressurised biomass gasification must be developed for large-scale plants, but first demonstration units will have to be built.
- For the use of biomass syngas in the FT synthesis, high liquid selectivity is desirable. The FT process (either a fixed bed or a slurry process) needs to be configured to fulfil this need. Both Shell and Sasol are performing much research on the FT synthesis.
- Development of sustainable forestry is necessary to ensure a large enough supply of clean wood. Efforts must be made to create a working biomass market and reduce prices of biomass, from various sources, over time.
- It should be investigated what the outlets are for FT naphtha and kerosene and if a premium for ‘green’ fuels will be paid for these products.

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1. Introduction

1.1 general background

To prevent climate change induced by human activity, green house gas emissions must be dramatically reduced. Renewable energy (e.g. solar, wind and biomass) could play a major role in achieving this. Biomass is a renewable energy source when carbon dioxide emissions caused by its use are absorbed by newly grown biomass. Biomass offers the possibility to produce liquid, carbon neutral transportation fuels. Ethanol, methanol, hydrogen and synthetic hydrocarbons can be produced from biomass and could offer feasible alternatives for the transport sector on foreseeable term (Williams et al. 1995). This is particularly relevant since transport is responsible for a large part of global CO_2 emissions. The global trend is that the share of transport in the total global energy consumption is increasing, especially in developing countries (Faaij et al. 2000). A recent study (Larson and Jin 1999) indicated that the use of Fischer-Tropsch (FT) technology for biomass conversion to synthetic hydrocarbons offers promising perspectives.

The Fischer-Tropsch process is a process capable of producing liquid fuels from syngas. The recent interest in Fischer-Tropsch (FT) synthesis has grown as a consequence of environmental demands, technological developments and changes in fossil energy reserves (Schulz 1999). First, FT liquids are totally free of sulphur and contain few aromatics compared to gasoline and diesel. Known reserves of natural gas have increased but a significant portion has been assigned 'stranded' (Agee 1998). Conversion on location into shippable hydrocarbon liquids is possible by FT synthesis. Products made by the FT synthesis, hydrocarbons of different length, can be transported by the same means as oil. Shell (natural gas based) and Sasol (coal based) apply FT synthesis on commercial scale. However, FT synthesis with *biomass* derived syngas has received little attention so far.

1.2 rationale

In principle, numerous process configurations for the conversion of biomass to FT liquids are possible, e.g. depending on gasifier types, the FT process considered and gas cleaning process. A scheme of the main process steps to convert biomass to FT liquids (and power) and possible different options is shown in figure 1.

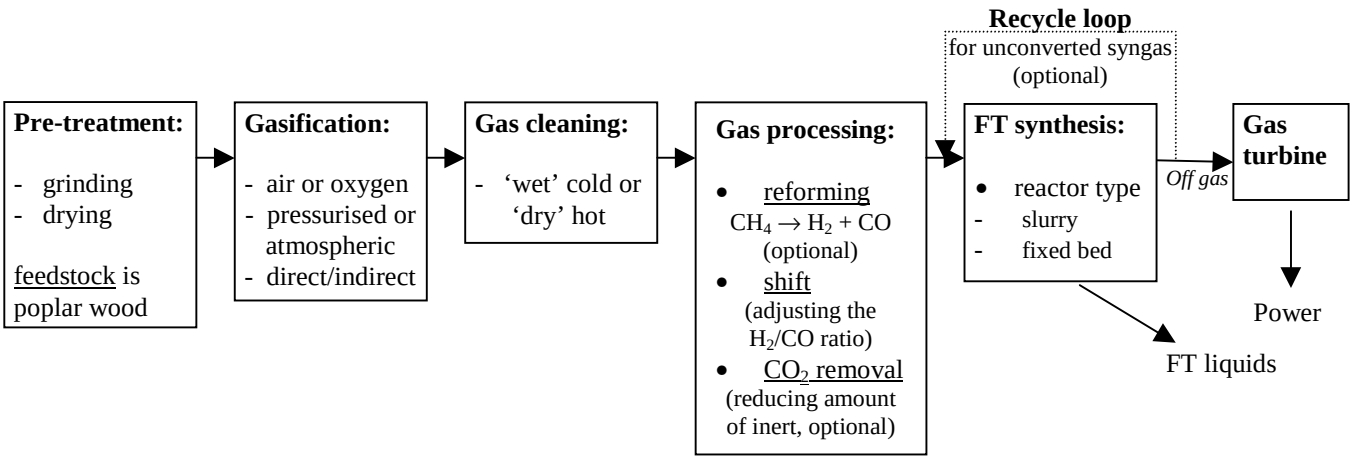

Figure 1: A schematic view of converting biomass to FT liquids combined with a gas turbine

Different gasification methods, covering atmospheric/pressurised, air-blown/oxygen-blown, indirect/direct gasification concepts, produce a wide range of syngas compositions, e.g. with H_2/CO ratios varying between 0.45 and 2 and different inert levels. Biomass derived syngas contains contaminants like H_2S, NH_3, dust and alkalis. The syngas is cleaned and processed to make it suitable for the Fischer Tropsch synthesis. Several processing steps can be applied to optimise the syngas composition for the FT reactor. The FT synthesis can be applied in different reactor types. Off gas from the FT synthesis can either be recycled partially (full conversion mode) or used directly for combustion in a gas turbine for the production of electricity (once through mode).

At present, biomass CFB (Circulating Fluidised Bed) gasification is used at relatively small scale (30-60 MWth) for co-firing and BIG/CC (Biomass Integrated Gasification Combined Cycle) installations. BIG/CC installations consist mainly of an atmospheric air blown biomass gasifier and a gas turbine. So, for these systems some 'real life' experience is gained. It still has to be proven if the gas cleaning section is able to meet the strict cleaning requirements for FT synthesis. Economies of scale are also a main issue.

1.3 objectives

The main objective of this study is to evaluate the different options to use biomass for the production of FT liquids and power. The main research questions are:

- to explore the technical feasibility and economics of BIG-FT processes in general
- to identify most promising system configurations; various biomass gasification processes will be studied in combination with FT-concepts in two main categories:
  3. Full conversion Fischer Tropsch with the possible use of a gas turbine, focussed on a maximum amount of FT liquids.
  4. Once through Fischer Tropsch with co-firing of the off gas in a gas turbine.

  To investigate economies of scale, the capacity considered of the BIG-FT systems is in the range of 100 to 1600 MWth.
- To identify key R&D issues essential for the commercialisation of BIG-FT technology.

1.4 methodology

The work consists of several steps. First of all a technology assessment on gasifiers, gas cleaning, syngas processing, FT conversion and combined cycles will be made to make an inventory of possible configurations. Besides information from literature, experts were consulted to identify the potential problems with the use of FT processes for biomass derived syngas. Manufacturers have been consulted for process data. The assessment includes some technologies that are not applied commercially at present.
Second, promising system configurations were selected for further performance modelling with help of the flowsheeting program Aspen plus. Aspen plus is used to calculate energy and mass balances.
Third, an economic evaluation is performed. Again, manufacturers have been consulted for cost data of various components. Fourth, an extensive sensitivity analysis is performed, including economies of scale of BIG-FT systems. Finally, the various system configurations are compared, conclusions drawn and recommendations on R&D issues are formulated.

2. System description

2.1 The Fischer Tropsch process

2.1.1 introduction and product description

The first Fischer-Tropsch process was developed in Germany about three quarters of a century ago. The FT synthesis produces hydrocarbons of variable chain length from a gas mixture of carbon monoxide and hydrogen. The synthesis can be performed in three main types of reactors: the fluidised bed reactor, the fixed bed reactor and the slurry phase reactor. Commercial scale FT plants have been installed and operated before and during World War II, mostly in Germany. Developments in the period shortly after World War II led to reactors suited for large-scale production of synthetic fuels. In the last 20 years there were several developments of more advanced reactors with (potentially) large capacity which are commercialised or ready for commercialisation. In 1993 Shell commissioned with their Shell Middle Distillate Synthesis (SMDS) process, a 12,000 bbl/day plant in Malaysia. Sasol developed the Sasol Slurry Bed Reactor (SSBR) technology, which is now considered for commercialisation (Sie and Krishna 1999).

The main mechanism of the Fischer Tropsch reaction is:

$CO + 2H_2 \rightarrow -CH_2- + H_2O \qquad \Delta H°_{FT} = -165 \ KJmol^{-1} \qquad (1)$

The $-CH_2-$ is a building stone for longer hydrocarbons. The various products of the FT synthesis (linear paraffins and α-olefins) and their trivial names are shown in table 1. The synthetic oil produced in the FT process can be distributed by the same infrastructure as conventional oil, like pipelines and ships.

Table 1:main products from FT synthesis and their trivial names (Laan 1999)

Chain length	Trivial name
C1-2	Fuel gas
C3-4	LPG
C5-12	Gasoline
C5-10	Naphtha
C11-13	Kerosene (Jet fuel)
C13-17	Diesel (Fuel oil)
C10-20	Middle distillates (Light gas oil)
C19-23	Soft wax
C24-35	Medium wax
C35+	Hard wax

A comprehensive description of all reactions in the FT synthesis is given in (Laan 1999) and in (Schulz 1999). The reaction takes place over a catalyst, either iron or cobalt[1]. Several schemes for the reaction mechanism have been presented (Schulz 1999), (Dry 1990/1991).

[1] Ruthenium is also used for experiments, but not on commercial scale because of it's high costs and limited world resources (Schulz 1999)

2.1.2 selectivity

A main characteristic regarding the performance of the Fischer Tropsch synthesis is the liquid selectivity of the process. Chain growth probability is determined by the chance a hydrocarbon chain either grows with another CH_2-group or terminates. The products made by the FT reaction are hydrocarbons of different length (besides a considerable amount of water). A high liquid selectivity is necessary to obtain a maximum amount of long (C_{5+}) hydrocarbon chains. At the same time the yield in the C_1-C_4 range decreases with increasing C_{5+} selectivity. Though the C_1-C_4 content in the off gas may be efficiently used in a gas turbine for power generation, maximum selectivity and conversion are favourable[2].

The relation between selectivity and the hydrocarbon yield is described by the Adams-Schulz-Flory (ASF) distribution (Schulz 1999), (Sie and Krishna 1999), (Larson and Jin 1999). The ASF distribution describes the molar yield in carbon number as: $\log C_n = \log \alpha^{n-1}(1-\alpha)$, where α is chain growth probability and n the length of the hydrocarbon. This means that α is the chance that the growth of a hydrocarbon continues with another CH_2-group, which makes $1-\alpha$ the chance the chain-growth terminates. Figure 2 shows the hydrocarbon yields for different values of chain growth probability.

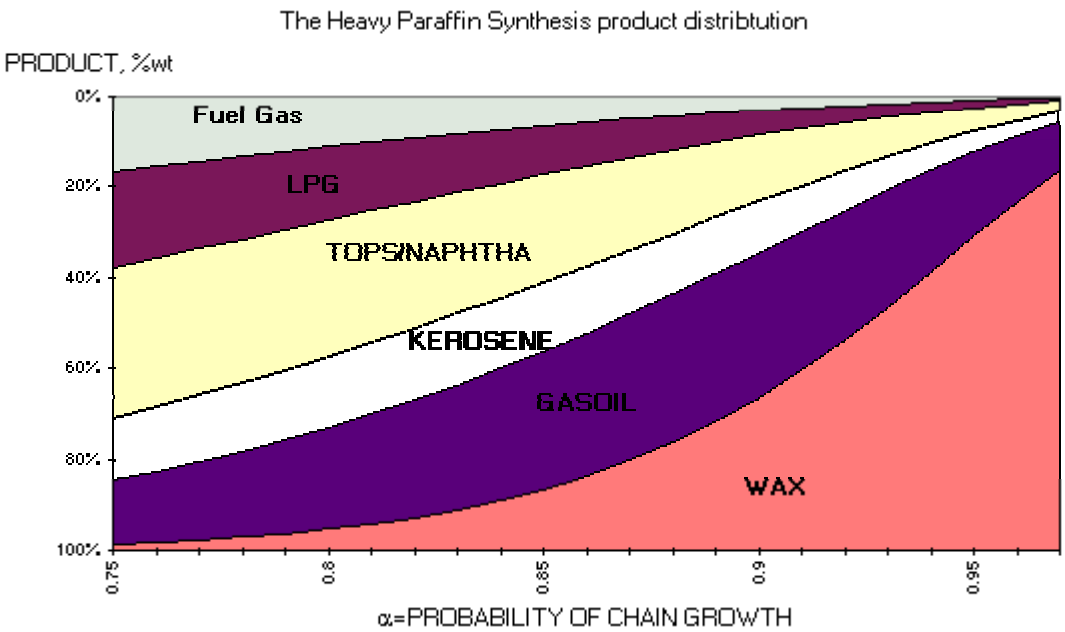

Figure 2: product distribution for different α (sww.shell.com 1999)

In practice, product distribution differs from the theoretical ASF-distribution (Geerlings, Wilson et al. 1999). This is mainly due to hydrogenolyse and reinsertion of olefins at a growth site. Hydrogenolyse is represented by the following reaction:

$C_n + H_2 \rightarrow C_{n-1} + CH_4$ (2)

Also, methanisation (3) takes place and a shift reaction (4):

$CO + 3H_2 \rightarrow CH_4 + H_2O$ (3)

$CO + H_2O \rightarrow CO_2 + H_2$ (4)

[2] It is more efficient to produce FT liquids (78% thermal efficiency) than power (approximately 55% efficiency using state-of-the-art combined cycles). If only C_{5+} is separated as liquid, it becomes more efficient to produce power if the C_{5+} selectivity drops below 35%.

This causes an increase in CH_4 and a decrease in C_{5+} selectivity. Other deviations from the ASF-distribution may be found in C_2 (lower than expected) and C_3+C_4 (higher). Also, at very higher carbon numbers, selectivity may slightly increase. In this paper α is used to describe the whole product range.
Selectivity is influenced by a number of factors, either catalyst dependent (type of metal (iron or cobalt), support, preparation, pre-conditioning and age of catalyst) or non-catalyst dependent (H_2/CO ratio in the feed gas, temperature, pressure and reactor type). Catalyst dependent influences on selectivity, apart from the division in iron and cobalt, are beyond the scope of this paper. The division in iron and cobalt is relevant since the water-gas-shift reaction is only significant over an iron catalyst (see 2.3.2). Also, iron catalysts are less expensive than cobalt. Regeneration or recovery of iron catalysts is not economically viable, while cobalt recovery is in most cases (Vosloo 2000).

- *H_2/CO ratio in the feed gas*

The FT synthesis uses H_2 and CO at a ratio depending on selectivity, when α=0 then H_2/CO=3, when α=1 then H_2/CO=2. For selectivity found in practise the H_2/CO ratio is close to 2.1. Gasification of biomass in most cases leads to a significantly lower H_2/CO ratio in the feed gas. Selectivity increases with decreasing H_2/CO ratio, because less H_2 is available to terminate chain growth. Chain termination takes predominantly place as an olefin (Hardeveld 1999). With decreasing H_2/CO, activity of the catalyst and therefore productivity decreases. By using a shift reaction (or by other reforming processes), adjustments can be made to the H_2/CO ratio.

- *Temperature and pressure*

The FT process is generally operated at pressures ranging from 20 to 40 bar. The temperature, depending on operation mode, ranges from 180 to 250 °C (low temperature Fischer Tropsch, LTFT) and from 300 to 350 °C (high temperature Fischer Tropsch, HTFT). Increasing temperature in the FT reactor results in a lower selectivity (Espinoza, Steynberg et al. 1999), (Dry 1981). As a result HTFT produces little wax and more gasoline where LTFT produces more waxes. Also, higher pressure leads to higher selectivity. So operating the FT reactor at optimal conditions is very important when a high selectivity is desired.

- *Inert*

The amount of inert in the syngas has negative influence on selectivity, because partial pressures of H_2 and CO are lowered. All main components in biomass derived syngas beside H_2 and CO (e.g. CH_4, N_2, and CO_2 [3]) can be considered inert for the FT synthesis. More inert in the syngas will cause the need for a higher space velocity[4] to obtain the same productivity. A higher space velocity will cause a large pressure drop.

- *Reactor type*

As mentioned earlier, there are three main kinds of FT reactors: the fluidised bed reactor, the fixed bed reactor and the slurry phase reactor. The two reactor types considered in this paper are the fixed bed and the slurry phase reactor. These two reactors are the most promising according to many authors e.g. (Sie and Krishna

[3] CO_2 is only inert for cobalt catalyst, since the water-gas-shift reaction takes place significantly over iron catalyst.
[4] Space velocity is the number of reactor volumes of feed at specified conditions that can be treated in unit time.

1999), some favouring the slurry phase reactor e.g.(Schulz 1999) and some favouring the fixed bed (Geerlings, Wilson et al. 1999).
The fluidised bed reactor is not being developed further. The main operational problem with the fluidised bed reactor is agglomeration[5] of the products inside the reactor and thus hampering fluidisation. Precautions preventing agglomeration will reduce the selectivity (Sie and Krishna 1999). Furthermore, fluidised beds are more expensive and mechanically complex (Vosloo 2000) and maintenance costs are high. A final disadvantage is that the separation of catalyst particles from the exhaust gas is problematic.

2.1.3 comparison between fixed bed and slurry Fischer Tropsch

A fixed bed reactor (figure 3) contains a large number of tubes. The catalyst is placed in the annular space between the tubes. Cooling water surrounding the outside of the tubes control the reactor temperature. Heat removal rates from a FT reactor is controlled by keeping the cooling medium water at the equilibrium vapour pressure at the desired temperature (Schulz 1999). The enthalpy of produced (low quality) steam can be used elsewhere in the process. Since conversion rates are rather low at less than 40% (Bechtel 1991-1994 quarterly reports), a recycle loop is generally used to recycle the unconverted gas and increase overall conversion. Recent developments however show higher once through conversion rates. Advanced fixed bed reactors are here assumed to have once through conversion up to 80%.

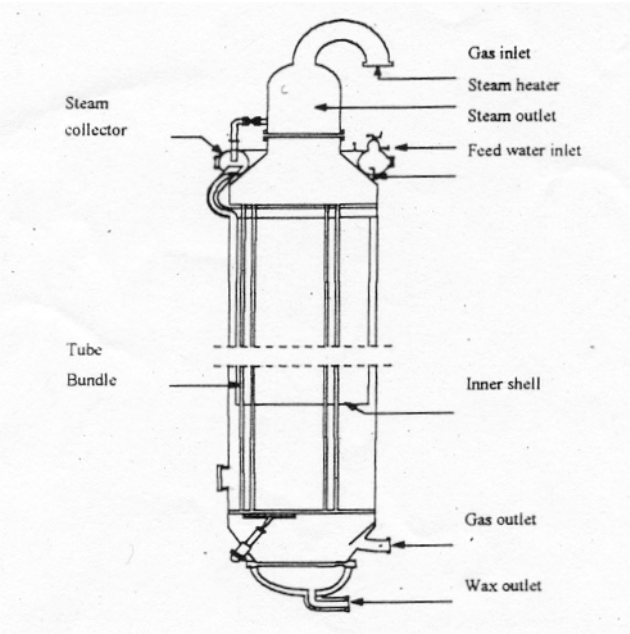

Figure 3: tubular fixed bed Fischer Tropsch reactor (Espinoza et al. 1999)

[5] Agglomeration is the sticking together of particles in the syngas. This leads to temperature inhomogeneity and eventually to shutdown.

A slurry reactor (figure 4) consists of a shell fitted with cooling coins in which steam is generated. Feed gas is distributed at the bottom and it rises through the slurry. The slurry consists of liquid reaction products, predominantly wax, with catalyst particles suspended in it. The reagent gases diffuse from the gas bubbles through the liquid phase to the suspended catalyst. The produced heavier hydrocarbons become part of the slurry, whereas the lighter gaseous products and water diffuse through the liquid to the gas outlet (Espinoza, Steynberg et al. 1999).

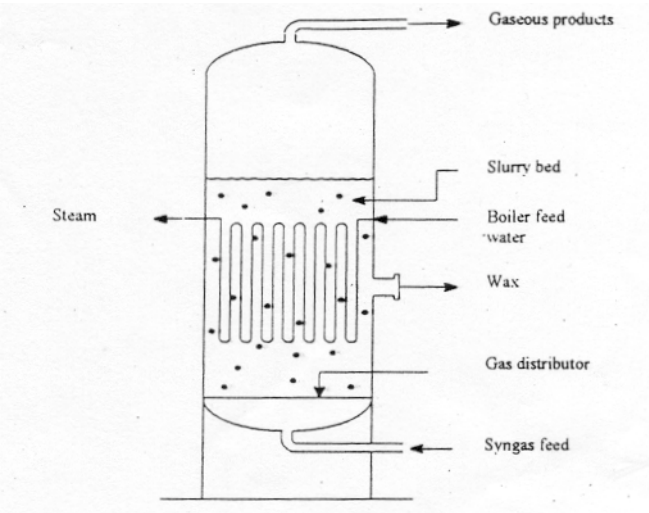

Figure 4: slurry bed Fischer Tropsch reactor (Espinoza et al. 1999)

Table 2 summarises the differences between the fixed bed and slurry phase reactors (Espinoza et al. 1999):

Table 2: differences between fixed bed and slurry

	Fixed bed	**Slurry**
Economic:		
Investment costs	High capital costs.	Less capital required.
O&M	Maintenance and labour intensive and long down time due to periodical catalyst replacement.	Little down time due to on-line catalyst replacement. Catalyst consumption is 66% less (Jager 1996) (Vosloo 2000).
Economies of scale	Scale up is straightforward, simply multiplying tubes, but no economies of scale obtainable. (Sie and Krishna 1999)	Possible, but scale-up is difficult. Laboratory studies are not representative for large scale plants (Laan 1999) (Sie and Krishna 1999).
Conversion efficiency:		
Once through conversion (depends much on H_2/CO ratio)	~ 40% (Bechtel 1991-1994 quarterly reports), up to 80% is assumed to be possible.	High average conversions (once through), up to about 80%. (Laan 1999) (Vosloo 2000)
C_{5+} selectivity	> 90% possible, negatively influenced by inert	> 90% possible, inert level is less critical.
Pressure drop	3-7 bar	< 1 bar (Jager 1997)
Technical aspects:		
Wax/catalyst separation	Performed easily and at low costs.	Main problem for commercial application. Sasol claims to have solved this properly (Vosloo 2000).
Process control	Difficult	Easy
Sensitivity to sulphur		1.5-2 times higher, so more thorough cleaning required.
Status:		
	Proven technology. Advanced reactors are likely to have higher once through conversion.	By some considered proven technology, but wax/catalyst separation is uncertain.

The competition between and further development of fixed bed and slurry reactors are ongoing. Specific biomass related advantages for either fixed bed or slurry cannot be given clearly, though sensitivity for inert (relevant for some biomass derived syngas compositions) seems less in a slurry reactor. Biggest disadvantage of the slurry reactor is the need for catalyst/wax separation, on which no public information appears to be available.

2.1.4 hydrocracking

When diesel is the desired final product, the FT product requires hydrocracking. Hydrogen is added to remove double bonds, after which the FT liquids are cracked

catalytically with hydrogen. Depending on the wax cracking conditions mainly diesel or kerosene is produced. The overall energy efficiency of the hydrocracking step is close to 100%. Also, hydrocracking conditions can be altered to obtain a desired product mix (table 3). If market prices should change a relatively simple adjustment to the hydrocracking process can be made to obtain the newly desired end products.

Table 3: Wax cracking conditions, wt%. On top some 4% gas (C_3 /C_4) is formed (Geijsel, 2000).

Product split	Gasoil mode	Kerosene mode
Naphtha	15%	25%
Kerosene	25%	50%
Gasoil (diesel)	60%	25%

The FT products are totally free of sulphur, nitrogen, nickel, vanadium, asphaltenes, and aromatics which all are typically found in mineral oil products. FT diesel, with a very high cetane number[6], can also be used as a blendstock to improve the quality of normal diesel. The FT naphtha has a much lower octane[7] number than 'normal' naphtha. FT kerosene for aviation still needs approval of several product specs. These products could therefore have less value than 'normal' naphtha and kerosene. But like FT diesel they contain no sulphur and other contaminants. Besides reducing emissions to air, this means FT liquids are probably well suited for use in fuel cell vehicles, which require very clean fuel to prevent damage to fuel cell catalyst.

[6] Cetane number is a primary measure of diesel fuel quality. It is essentially a measure of the delay before ignition. The shorter the delay the better – and the higher the cetane number

[7] Octane number is a quality rating for gasoline, indicating the ability of the fuel to resist premature detonation and to burn evenly when exposed to heat and pressure in an internal combustion engine. Normal gasoline has a octane number of 87-89, FT gasoline has a very low octane number.

2.2 Biomass gasification

2.2.1 pre-treatment of feedstock

A wide variety of biomass can be used as feedstock. Wood, agricultural wastes, organic wastes, and sludges are each potential fuels (Faaij, 1997). However, in this study clean (poplar) wood is used as feedstock. Clean wood gives a relatively clean syngas, with low levels of contaminants. On the longer term wood from dedicated plantations may be a major source of renewable biomass (see e.g. Hoogwijk et al 2000).

Pre-treatment prior to gasification is required and generally consists of screening, size reduction, magnetic separation, 'wet' storage, drying and 'dry' storage (Faaij et al. 1998). Moisture content of the 'wet' poplar chips delivered is assumed to be 30% (Faaij al. 1998) (Katofsky 1993) (Consonni and Larson 1994). Drying is generally the most important pre-treatment operation, necessary for high cold gas efficiency (Faaij 1997) gasification or combustion. Drying reduces the moisture content to 10-15%. Drying can either be done with flue gas or with steam. Since a lot of low quality steam is generated in the FT process, steam drying is preferable (e.g. a Niro steam dryer, see (Pierik 1995)). Also, steam drying results in (very) low emissions and may eventually be safer with respect to risks for dust explosion. Per tonne dry biomass input 0.41 tonne steam (±12 bar, ±290°C) is needed for drying to 10% moisture content, 0.33 for drying to 15% moisture content (Pierik 1995).

2.2.2 gasification

Conversion of biomass to an H_2 and CO containing feed gas that is appropriate for FT synthesis takes places through biomass gasification. Gasification can take place at different pressures, either directly heated or indirectly heated (lower temperatures), and with oxygen or air. Direct heating occurs by partial oxidation of the feedstock, while indirect heating occurs through a heat exchange mechanism. Fixed bed gasifiers, either co- or countercurrent, have a power range from 2-6 MWth. Since the investigated range is 100-1600 MWth, the biomass gasifiers that have been taken in consideration are fluidised bed gasifiers. Advantages and disadvantages of each gasification method are shown in table 4.

Table 4: main technical aspects gasification method

Pressurised + pressurised equipment upstream is smaller and generally more economical at larger scales (see 4.1) - higher costs of gasifier at small scale - high risk in keeping constant mass flow in gasifier, operational experience so far limited to demonstration projects (Rajani 2000)	**Atmospheric** - larger equipment upstream needed + less costs at small scale (see 4.1) + significant commercial experience
Oxygen - air separation plant needed, especially at small scales relatively expensive + no dilution of syngas by N_2	**Air** + cheaper - N_2 diluted syngas, has negative influence on C_{5+} selectivity - larger equipment needed upstream
Direct + less tars produced; the presence of tars in the syngas is one of the biggest problems when gasifying biomass	**Indirect** + no N_2 dilution even if air is used - bigger tar problem -> currently demonstrated (BCL)

To cover the whole range of gasification processes, different gasifiers currently available and/or under development have been selected (table 5). These gasifiers produce a wide range of syngas compositions (table 6).

Table 5: gasifiers included in this analysis

gasifier		process type
BCL	= Batelle Colombus gasifier	indirect, air blown, atmospheric
IGT	= Institute of Gas Technology gasifier	direct, oxygen, pressurised
IGT+	= idem, with process adjustments	direct, oxygen, pressurised
EP	= Enviro Power gasifier	direct, air blown, pressurised
TPS	= TPS gasifier	direct, air blown, atmospheric

Table 6: operating characteristics of the gasifiers evaluated in this paper, based on poplar wood.

Gasifier type: / Characteristics:	BCL (Williams, 1994)	IGT (Katofsky 1993)	IGT+ (estimates) (Knight 1999)	EP (with dolomite tar cracker) (Faaij, Ree et al. 1998)	TPS (with dolomite tar cracker) (Lassing et al. 1995)
P (bar)	2*	34	20.3	22	1.3
T (K), exit	1136	1255	1241	1223	1173
Moisture dry biomass	10%	15%	15%	15%	15%
Pilot size (dry tons/day)	10	100 (Knight 1999)	No pilot	-	270
Flowrate dolomite (kg/kg wet)	-	0	0	0.0095	0.0268
Flowrate air/oxygen[8]	1.46 kg/kg dry	0.3 kg/kg dry	0.3 kg/kg dry	-	1.4 kg/kg wet
Steam (kg/kg wet input)	0.19	0.34	0.6	0,34 *	0.34 *
Yield (kmol/ dry tonne)	45.8	82.0	123.1	113.3	112.1
LHV syngas (MJ/Nm^3 wet gas)	13.9	7.3	4.8	5.8	5.2
Gasifier efficiency[9]	86.8	80.7	80.9	88.6	80.0
H_2/CO ratio	0.45	1.39	2.0	0.73	0.77
Composition (mol % [dry])					
H_2O	19.9 [0]	31.8 [0]	50.6[0]	13.55*[0]	13.55[0]
H_2	16.7 [20.8]	20.8 [30.5]	15.68[31.7]	10.03[11.6]	13.25[15.3]
CO	37.1 [46.3]	15.0 [22.0]	7.83[15.85]	13.83[16.0]	17.22[19.9]
CO_2	8.9 [11.1]	23.9 [35.0]	17.71[35.9]	15.4[17.8]	12.22[14.1]
CH_4	12.6 [15.7]	8.2 [12.0]	5.73[11.6]	7.26[8.4]	2.82[3.26
C_{2+}	4.8 [6.0]	0.3 [0.5]	0	0.48[0.62]	0.96[1.11]
C_2H_4	4.2 [5.2]				0.94
C_2H_6	0.6 [0.74]				0.02
N_2+Ar	0	0	0.40[0.8]	38.9[45.3]	39.20[45.3]
others	<0.3	<0.3	<0.3	<0.3	<0.3

* =assumption
\- =not available

The syngas produced by the different gasifiers contain various contaminants: particulates, condensable tars, alkali compounds, H_2S, HCl, NH_3, HCN and COS (Ree, Oudhuis et al. 1995). Remarkably, no full data sets of syngas compositions including these contaminants are available for the gasifiers considered.

2.3 Syngas processing

The syngas, produced by the gasification of biomass, consists mainly of H_2, CO, CO_2 and CH_4. Their shares in the syngas can be tailored to the FT process by syngas processing. Syngas processing consists of three steps, which are all optional.

2.3.1 methane reforming

The syngas can contain a considerable amount of methane (CH_4) and other light hydrocarbons, which can represent a significant part of the heating value of the gas. Methane reforming converts methane and other light hydrocarbons (with steam) to H_2

[8] Oxygen of 99.5% purity is used; production is assumed to require 305 kWh per tonne oxygen for 95% purity. This is somehow scale-dependent (Air products 1991) Oxygen of 99.5% purity will require 15% extra energy (Williams, 1994).
[9] Gasifier efficiency is defined as [energy content syngas/energy content biomass input], based on LHV. Energy content of steam and air/oxygen added is not taken into account.

and CO. In principle various reforming processes (e.g. Autothermal Reforming (ATR), partial oxidation (POX) and steam reforming (SMR)) can be considered. However, ATR (with either air or oxygen) proved to be the only method applicable for the used syngas compositions (see appendix 2). POX and SMR are unsuited due to high CO and C_{2+} content of the gas, causing Boudouard reactions[10] and carbon deposition as a result of spontaneous cracking (Grootveld, 2000) (Huisman, 2000). This would destroy the reformer.

In the ATR, part of the feed is oxidised to produce CO_2 and H_2O, sufficient to supply the necessary heat to reform the remaining feedstock. When oxygen is available (as is the case when the IGT gasifier is used) this is also used for the ATR, because advantages of scale lower the relatively high costs of oxygen production. When oxygen is not present air can be used, but this adds extra inert into the syngas.

Autothermal reforming is represented by the following reactions:

$CH_4 + H_2O \leftrightarrow CO + 3H_2$ $\Delta H° = +206$ MJ/kmol (5)

$C_2H_4 + 2H_2O \leftrightarrow 2CO + 4H_2$ $\Delta H° = +210$ MJ/kmol (6)

$C_2H_6 + 2H_2O \leftrightarrow 2CO + 5H_2$ $\Delta H° = +347$ MJ/kmol (7)

$CH_4 + 2O_2 \leftrightarrow CO_2 + 2H_2O$ $\Delta H° = -636$ MJ/kmol (8)

The ATR operates at 900°C and at pressures up to 70 bar and is close to commercialisation.

2.3.2 the water-gas shift reaction

The syngas produced by the gasifier generally has an H_2/CO ratio lower than 2, which is the required H_2/CO ratio for FT synthesis with a cobalt catalyst. The water-gas shift reaction converts CO into H_2 (with steam) and thereby raises the H_2/CO ratio.

$CO + H_2O \leftrightarrow CO_2 + H_2$ $\Delta H°_{WGS} = -41.3$ $KJmol^{-1}$ (9)

The shift reactor operates at 330 °C and at 15-25 bar.

When using an iron catalyst in the FT reactor, the shift reaction takes place in the FT reactor (Laan 1999). Combining the WGS reaction and the FT reaction results in the overall reaction:

$2CO + H_2 \leftrightarrow -CH_2- + CO_2$ (10)

When a cobalt catalyst is used, the shift reaction has to be performed with an external shift reactor. The extra costs are relatively low compared to other investment costs (1-2%). So the difference between using an iron or a cobalt catalyst in this respect is rather small.

2.3.3 CO_2 removal

The syngas produced by the various gasifiers contain considerable shares of CO_2. When methane is reformed or when a shift reaction is applied CO_2 content will increase further. CO_2 is considered inert in the FT process and thus has negative influence on C_{5+} selectivity. Inert will build-up when a recycle loop is used.

Making a choice between the solvent absorption processes available for CO_2 removal is mainly based on partial pressures of CO_2. At high partial pressures, physical solvent processes are most attractive. However, at partial pressures of CO_2 smaller than 10 bar, Sulfinol is the best method for CO_2 removal (Klinkenbijl, 2000). Sulfinol, a combined physical and chemical solvent process, has positive effects on the required cleaning upstream. Sulfinol D in combination with FT has the additional benefit that the necessary COS removal takes place as well. It operates at 10°C and at pressures up to 70 bar. More than 98% of CO_2 in the gas is removed.

[10] A good description of the carbon deposition process, for different C:H:O ratios and temperatures and pressures, is given in (Katofsky 1993)

2.4 Power generation with a combined cycle

2.4.1 Gas turbine

The off gas from the FT reactor can be used in a gas turbine to generate electricity. The gas turbine consists of two distinct parts. First the off gas, at high pressure, is put through a combustor. Adding pressurised air generates a large hot stream, which is expanded so electricity is generated. Part of the work is used to drive the air compressor. In some cases the syngas has a caloric value too low to ensure stable combustion in the combustor. The mass flow will become too large if the standard combustion temperature is to be maintained. Therefore, a lower combustion temperature is necessary when syngas with a low caloric value is put through a gas turbine, also adjustments to the gas turbine have to be made (Ree, Oudhuis et al. 1995). This reduction of gas turbine combustion temperature is called de-rating. Another form of de-rating is raising the pressure of the air put through the turbine. A recent study by (Walter et al. 1998) shows that moderate de-rating is allowable for some gas turbine models. Heavy de-rating has strong impacts on overall performance and economics of combined cycles. Roughly spoken, a syngas with LHV higher than 6 MJ/Nm^3 can be burned without de-rating, at 1200°C and 14 bar. The burner may need modifications if the syngas has LHV lower than 10 MJ/Nm^3. (Walter et al. 1998). Syngas with a caloric value between below 6 MJ/Nm^3 will be burned at 1100 °C and 16 bar. Syngas with a caloric value below 5 MJ/Nm^3 will probably not be suited for a gas turbine, but this depends on the gas turbine used (LHV's of 2.5 have been used in gas turbines). A possible option for using this very low caloric value syngas is co-firing of methane to raise the caloric value of the mixture. Co-firing has a large potential and will increase efficiency. Co-firing the off gas in a natural gas based gas turbine is an option to secure a constant electricity generation in the gas turbine. If the biomass gasifier fails, the gas turbine can operate entirely on natural gas. If the FT part fails or if catalyst replacement has to be performed, this section can still be operated. In such a case the syngas is put directly into the gas turbine, so it operates (almost) entirely on biomass derived syngas.

When a gas turbine is used the exhaust gas carries a large amount of heat. Steam from cooling the exhaust gas is fed to the Heat Recovery Steam Generation (HRSG). The exhaust gas of the gas turbine cannot be cooled below 170°C due to environmental restraints (Faaij 1997). A gas turbine using HRSG is called a combined cycle.

2.4.2 Heat Recovery Steam Generation (HRSG)

At different places in the whole process cooling down is necessary, e.g. the exhaust gas from the gas turbine, the cooling down of the syngas after gasification and the FT liquid separation. Water is pressurised in advance of the heat exchanger, so high-pressure steam is generated (70 bar, 500°C or 41 bar, 440°C). The superheated steam is expanded in a partly condensing gas turbine to produce electricity. Also, steam required in the process (drying, gasification, shifting and reforming) can be taken from the HRSG at different pressures.

2.5 Gas cleaning

2.5.1 general

The syngas produced by the gasification process contains different kinds of contaminants, viz. particulates, condensable tars, alkali compounds, H_2S, HCl, NH_3 and HCN (Ree, Oudhuis et al. 1995). These contaminants can lower activity in the FT synthesis by poisoning the catalyst. Sulphur is an irreversible poison for the FT catalyst (and to a lesser extent for the shift and reformer catalyst), because it will stick to the former active site. Tolerance is low and 'deep' cleaning is required. Two distinct routes of cleaning will be considered, viz. 'wet' low temperature cleaning and 'dry' hot temperature cleaning.

2.5.2 conventional 'wet' low temperature cleaning

Conventional 'wet' low temperature cleaning (figure 5), as described in (Ree, Oudhuis et al. 1995) and (Faaij 1997), is being proposed to clean the fuel gas for BIG/CC installations. However, cleaning requirements for the FT synthesis are much more stringent as for the BIG/CC systems.

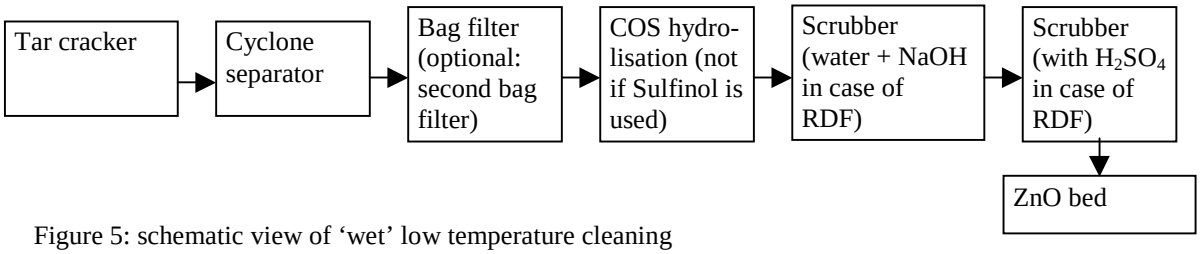

Figure 5: schematic view of 'wet' low temperature cleaning

Specifications for FT are given in table 7 and compared to syngas compositions typical for clean wood. Since relatively clean natural gas is the common feedstock for FT synthesis, actual cleaning specifications are not known for some specific biomass contaminants. Therefore the given specifications are estimates. More information could be obtained from coal based FT plants, since coal also contains high levels of contaminants. The specification for sulphur however is explicitly known, since it is also present in natural gas and a known irreversible poison.

Table 7: contaminant concentrations (wt%) and their maximum values for FT synthesis (ppb)

Contaminant present in dry feedstock (in syngas)	**poplar wood** adapted from (Katofsky 1993)	**Assumed cleaning requirements FT in ppb**	**Cleaning efficiency required**	**Required cleaning steps ('wet' low temperature cleaning)**	**Results + = requirement achieved ? = some uncertainty**
Ash (particulates)	1.33	0	>99.9%	Cyclone separator, bag filters/scrubber.	+?, small particles may be problematic
N (HCN+ NH_3)	0.47	20	>99.9%	Scrubber (possibly with H_2SO_4), Sulfinol D also removes HCN and NH_3.	+?
S (H_2S+COS)	0.01	10	>99.9%	Scrubber, possibly COS hydrolisation unit or Sulfinol D necessary, ZnO guard bed.	++, ZnO guard beds are also used for natural gas based FT
Alkalis	0.1*	10	>99.9%	During cooling down alkalis condense on particulates, possibly also on vessels (and thereby polluting them).	+?
Cl (HCl)	0.1	10	>99.9%	Absorbed by dolomite in tar cracker (if used), reaction with particulates in bag filter, scrubber (possibly with NaOH).	+
Pb and Cu	0*	not known	-	Condense on particulates, but actual behaviour has not been studied.	?
Tars	-	0	>99.9%	Condense on particulates and vessels (and thereby polluting them) when syngas is cooled below 500°C.	??, potential tar problem, limited experience with complete removal or conversion for biomass .

*=adapted from (Faaij 1997) for miscanthus
- not known, but order of magnitude is g/Nm^3

As can be seen from table 9, it cannot be said with certainty that all cleaning specifications will be met by 'wet' low temperature cleaning. But potential problems with tars give rise to discussion and further research is required.

2.5.3 advanced 'dry' hot gas cleaning

Hot gas cleaning consists of several filters and separate units in which the syngas can be cleaned at high temperature. So the high temperature of the syngas can (partly) be maintained, resulting in an efficiency advantage. Hot gas cleaning is specifically advantageous when a reformer or shift reactor is applied directly after the cleaning section, because of their high inlet temperatures. When FT is applied directly after hot gas cleaning the syngas has to be cooled to 200 °C anyway. For atmospheric gasification hot gas cleaning does not improve efficiency, because cooling is required before the syngas is compressed. Compressing at higher temperature results in an unacceptably high energy demand of the compressor.

Hot gas cleaning is not a commercially demonstrated process yet; some unit operations are still in the experimental phase. Disadvantageous for the application of hot gas cleaning in this case is the high purification requirements of FT synthesis (see table 7). It is highly questionable if hot gas cleaning will meet these standards in a foreseeable timeframe. Research is so far aimed at developing hot gas cleaning to

meet the requirements for BIG/CC installations for which fuel gas requirements are less severe. So far there are no commercial-scale processes for the high temperature removal of nitrogen compounds, halides, alkali metals and heavy metals (Mitchell 1998). Contaminants are separated mostly as salts that can easily be processed and possibly utilised.

2.5.4 discussion

'Wet' low-temperature fuel gas cleaning is the preferred on the short term (Ree et al. 1995). Conventional 'wet' low-temperature cleaning will have some thermodynamic penalty and will require additional waste water treatment but there is little uncertainty at present as to the cleaning effectiveness of such systems, with respect to BIG/CC installations (Consonni and Larson 1994). However, actual testing with biomass/FT systems is necessary to ensure the effectiveness of 'wet' low temperature cleaning for these systems.

Within 10 years hot gas cleaning may become commercially available for BIG/CC installations (Mitchell 2000), but opinions differ. Even when BIG/CC requirements are met, significant improvements are necessary to meet the much more severe FT requirements. As a result, hot gas cleaning is considered as a (very) advanced option. To show the possible effect of hot gas cleaning on the performance of BIG-FT systems, it has been included in the modelling work.

2.6 System considerations and selection

2.6.1 main selection

In theory, a large number of system configurations to convert biomass to FT liquids and power is possible (table 8).

Table 8: possible system configurations for BIG-FT

Gasifier	reforming	shift	FT system	FT data
• BCL • IGT • IGT+ • EP • TPS	• ATR • no ATR	• intern (iron) • extern (cobalt)	• full conversion once through • fixed bed slurry	1. α 2. once through efficiency

At this point a selection is made. The concepts are divided into two main groups:

1. Full conversion Fischer Tropsch with the possible use of a gas turbine, focussed on a maximum amount of FT liquids.
2. Once through Fischer Tropsch with co-firing of the off gas in a 150 MW -natural gas based- gas turbine.

For all concepts an external shift reactor will be used, implying the use of a cobalt catalyst. This will have little effect on overall outcome, as argued in 2.3.2. Since it is difficult to predict a specific liquid selectivity for biomass derived syngas, with different inert percentages and possibly unexpected contaminants, a realistic range of α=0.8 to 0.9 is modelled. When α is 0.8 this means 73.7% of the converted CO is used to form C_{5+}, when α is 0.9 this is 91.9%. As a conventional option a once through CO conversion of 40% is modelled for the full conversion concepts. The

relatively low once through CO conversion makes a recycle loop very useful to increase overall CO conversion. As an advanced option 60% and 80% once through CO conversion is modelled for the once through concepts. Advanced FT reactors, either fixed bed or slurry are assumed to obtain higher once through CO conversion.

2.6.2 full conversion Fischer Tropsch, with possible use of a gas turbine

The goal of the full conversion concepts is a focus on FT liquids and to a lesser extent on power production. Modelling is done with and without ATR, because ATR can increase the amount of FT liquids produced. The IGT plus option is modelled without ATR and shift, otherwise the basic idea of creating a syngas with a 'perfect' H_2/CO ratio would not make sense. In case of the TPS-based systems, ATR with air will lower the H_2 and CO content in the syngas because some H_2 and CO has to be burned to obtain the necessary heat. Therefore, this configuration is not modelled.
Sulfinol D is used for CO_2 removal for all concepts. Less inert will allow for a high C_{5+}-selectivity and will therefore increase the amount of FT liquids produced. Sulfinol D does not have the same effect for each gasifier. The amount of inert in the syngas of the IGT gasifier is reduced from 36% to 1%, because of the high CO_2 content. For the TPS gasifier, the amount of inert is reduced from 79% to 59% (because N_2 remains in the syngas).
A gas turbine has been modelled if the LHV of the off gas is sufficiently high. If the LHV of the off gas drops below 5 MJ/Nm^3 the concept involved is modelled both with and without a gas turbine.

2.6.3 once through Fischer Tropsch, with co-firing in a 150 MW natural gas based gas turbine

The once through concepts are less focussed on FT liquids and more on power. All gasifiers are modelled without ATR and Sulfinol D, so no optimisation for the FT synthesis is made. The IGT gasifiers are also modelled with Sulfinol D, due to their high CO_2 content of the syngas. The co-firing with natural gas of the gas turbine results in higher electrical efficiency of the combined cycle due to scale effects that (also) result in higher efficiency.

3. System calculations

3.1 Modelling principles

Modelling of the various concepts has been performed with the flowsheeting program Aspen plus, on basis of the flowsheet presented in figure 6. The energy and mass balances of the gasifier products (table 7, 2.2.2) have been used as starting point for the calculations (gasifiers are not modelled). Scale used for the calculations is 427 MWth HHV or 367 MWth LHV (equivalent to 80 tonne dry biomass per hour).

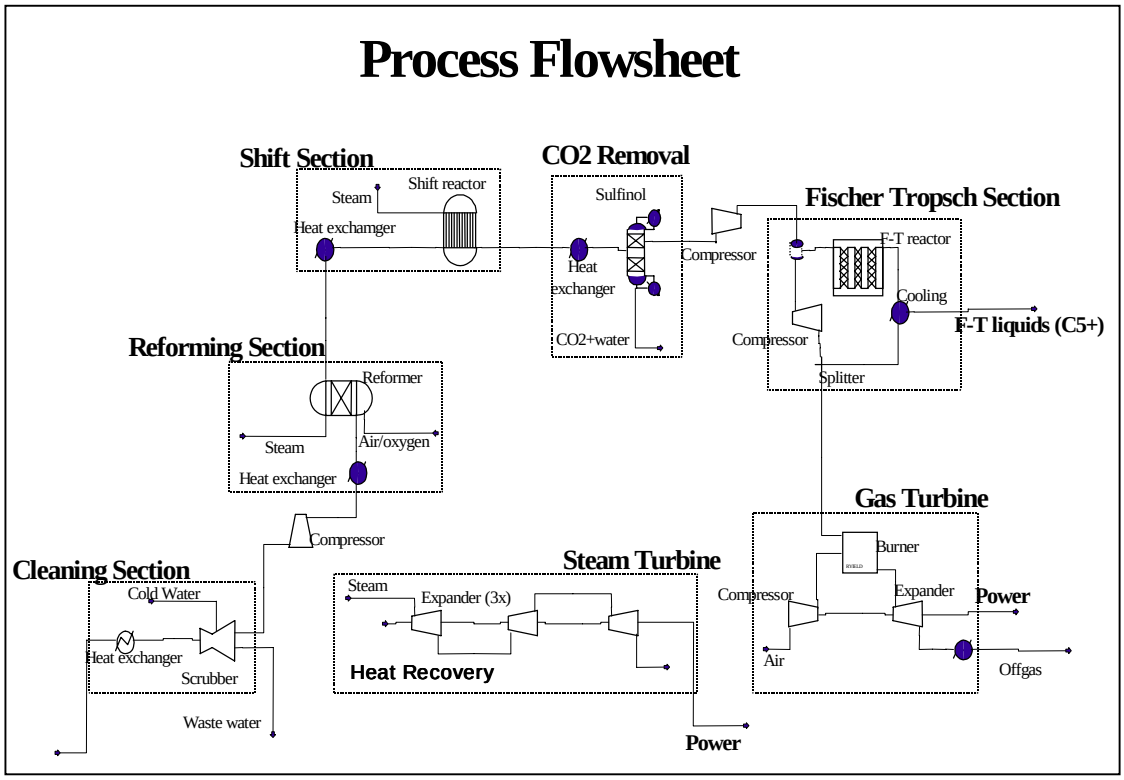

Figure 6: Aspen flowsheet used for the calculation of the mass and energy balances

When 'wet' low temperature gas cleaning is used, the inlet temperature of the scrubber is 400°C, the outlet temperature of the scrubber is 40°C below the dew point (depends on pressure). In case of 'dry' hot gas cleaning the outlet temperature is 450°C.

Heat integration is done without a pinch analysis (detailed heat integration), but by means of matching heat supply and demand. Temperature difference between hot stream and cold stream has a minimum of 15°C. Syngas/syngas heat exchangers (necessary in case ATR is used) cannot heat the cold stream above 400°C, due to coking problems (Huisman 2000). Syngas/steam heat exchangers have no practical temperature limit. An overall energy efficiency for the steam turbine of 72% is used. Steam from the HRSG is used to produce steam for biomass drying, biomass gasification, reforming and shifting. The outlet pressure of the steam from the HRSG is 0.04 bar, at this pressure the steam is at environmental temperature.

The FT recycle loop is modelled with a 2:1 recycle:fresh feed mol ratio. In practice a recycle loop could be replaced by two FT reactors in line, to avoid inert build-up. The H_2/CO ratio after the shift reaction is 2.0. The effective outlet H_2/CO ratio was assumed to be >0.4 After the FT synthesis, cooling to below 50°C makes it possible for the C_{5+} fraction to be separated as liquid. Further cooling to separate C_3 and C_4

could be done, but whether this is economically viable is unknown. In the concepts modelled, C_1-C_4 is used in the gas turbine. Details about Aspen modelling are given in appendix 4.

3.2 Results

The overall energy efficiencies[11] of the ***full conversion concepts*** are presented in table 9, with a division of outputs in FT liquids and net power production (or use). Details are given in appendix 5.

Table 9: overall energy efficiencies (LHV) of the full conversion concepts. α=0.8: C_{5+}=73.7%; α=0.85: C_{5+}=83.5%;α=0.9: C_{5+}=91.1%

concept (1)	BCL R	BCL R-nt	BCL	IGT R	IGT R-hg	IGT	IGT+	EP R	EP R-nt	EP	TPS
α=0.80 eff. %	45.1	30.1	35.9	44.7	46.1	46.0	44.6	41.6	25.4	42.4	32.9
FT-liquids	123.9	123.9	66.8	129.9	132.6	81.1	78.4	100	100	64.6	83.1
Power	41.4	-13.6	65.0	34.2	36.6	87.5	85.4	52.5	-6.7	90.9	37.6
α=0.85 eff. %	47.0	33.5	37.1	47.7	49.1	47.4	44.9	44.8	29.4	43.4	34.5
FT-liquids	139.1	139.1	75.9	150.7	153.4	91.9	89.8	113.3	113.3	73.2	94.2
Power	33.3	-16.3	60.4	24.4	26.8	82.2	75.0	50.9	-5.5	86.2	32.5
α=0.90 eff. %	48.0	38.2	38.1	50.1	51.5	48.2	47.3	45.4	32.2	44.5	35.8
FT-liquids	154.8	154.8	83.4	168.7	171.4	101.7	97.9	124.6	124.6	80.5	103.6
Power	21.3	-14.6	56.5	15.2	17.4	75.1	75.7	42.0	-6.5	82.9	27.6

(1) Explanation of used codes:
BCL, IGT, IGT+, EP and TPS= gasifier names,
R= reformer used,
nt= no gas turbine,
hg= hot gas cleaning

The BCL-R and the EP-R concepts produce off gas with a LHV< 5 MJ/Nm^3, which is probably too low for (direct) use in a gas turbine. The gas turbine is of particular importance for those concepts because without the gas turbine, a substantial amount of energy, in the form of light hydrocarbons (C_1-C_4) or compressed N_2, would be wasted. Co-firing with natural gas could be necessary for these concepts, to upgrade the LHV of the off gas.

The overall energy efficiencies of the ***once through concepts*** are presented in table 10 and 11. All once through concepts make use of a 150 MW gas turbine, the net power output in the table however is based on the share of the off gas.

[11] The overall energy efficiency of the systems is defined as:
sum of all outputs / total biomass input.

Table 10: overall energy efficiencies (LHV) of the once through concepts, 60% conversion. α=0.8: C_{5+}=73.7%; α=0.85: C_{5+}=83.5%;α=0.9: C_{5+}=91.1%

concept (1)	BCL	IGT	IGT-S	IGT+	IGT+-S	EP	TPS
α=0.80 eff. %	35.8	44.1	43.9	43.5	43.6	43.6	33.9
FT-liquids	49.9	58.9	58.9	57.8	57.8	54.8	69.3
Power	81.3	103.0	102.2	101.9	102.4	105.1	55.2
α=0.85 eff. %	37.2	44.6	44.5	44.6	44.7	44.6	35.2
FT-liquids	56.6	66.8	66.8	65.4	65.4	62.0	78.5
Power	79.8	97.0	96.4	98.1	98.8	101.8	50.7
α=0.90 eff %.	38.0	45.6	45.4	45.4	45.7	45.5	36.3
FT-liquids	62.2	73.5	73.5	72.0	72.0	72.0	86.3
Power	77.2	93.7	93.1	95.0	95.6	98.6	47.0

(1) *Explanation of used codes: S= Sulfinol D used*

Table 11: overall energy efficiencies (LHV) of the once through concepts, 80% conversion. α=0.8: C_{5+}=73.7%; α=0.85: C_{5+}=83.5%;α=0.9: C_{5+}=91.1%

concept	BCL	IGT	IGT-S	IGT+	IGT+-S	EP	TPS
α=0.80 eff. %	37.7	46.4	46.2	45.6	45.7	47.2	36.5
FT-liquids	66.6	78.6	78.6	77.0	77.0	73.0	92.4
Power	71.9	91.5	91.0	90.4	90.8	100.4	41.5
α=0.85 eff. %	38.9	47.8	47.7	47.0	47.1	48.6	38.2
FT-liquids	75.4	89.1	89.1	87.3	87.3	82.7	104.6
Power	67.4	86.3	85.9	85.3	85.8	95.7	35.6
α=0.90 eff. %	40.4	49.0	48.9	48.2	48.3	49.8	39.7
FT-liquids	82.9	98.0	98.0	96.0	96.0	91.0	115.1
Power	65.3	81.9	81.6	81.0	81.4	91.6	30.5

3.3 Discussion of results

The main energy losses in the model are on the account of the gasification section (approximately 80% thermal efficiency), the shifting and reforming section (approximately 10% energy loss), the FT section (78% thermal efficiency for the converted CO and H_2) and the combined cycle (approximately 50% efficiency). Important is that the small hydrocarbons (C_1-C_4) formed in the FT synthesis and used in the gas turbine, have an energy loss both in the FT section as in the combined cycle. Also, from the FT section (35 bar outlet pressure) to the gas turbine (14-16 bar inlet pressure) the syngas loses pressure without producing electricity. An expansion turbine could possibly be used to reduce this energy loss

The modelled concepts give the following insights:

1. CO_2 removal (Sulfinol D) has little effect on overall energy efficiency (but does have an effect on the amount of inert and thus on liquid selectivity). Question remains if the investment is justified by the improvement of selectivity.
2. Higher C_{5+} selectivity leads to higher overall energy efficiency for all concepts. The C_{5+} selectivity that is obtainable is uncertain, but will be in the given range. When much inert is present (as is the case with air gasification), C_{5+} selectivity will probably be in the bottom part of the used range. When little inert is present C_{5+} selectivity will be higher.
3. The concepts with 80% conversion have higher energy efficiencies than the concepts with 60% conversion. So a high overall CO conversion has a beneficial effect on efficiency.
4. Pressurised concepts have higher overall energy efficiencies than atmospheric

concepts. The difference is about 10 percent points. This is mainly due too the high electricity consumption of the syngas compressors when atmospheric gasifiers are used.

5. Methane reforming does only improve efficiency when C_{5+} selectivity is high; these concepts are most sensitive to C_{5+} selectivity. For the BCL concepts methane reforming does also improve efficiency at low C_{5+} selectivity, due to the high C_{2+} content of the syngas.
6. Hot gas cleaning improves efficiency (for pressurised concepts) with 1.4 percent points.

The concepts with high overall energy efficiency are based on pressurised gasifiers. When high C_{5+} selectivity (α=0.90) is assumed, IGT-R has an LHV efficiency of 50.1%, when hot gas cleaning is used 51.5%. The 80% once through concepts show that efficiencies of near 50% are obtainable for the EP and the IGT gasifiers, even without the use of a reformer.

4. Economics

4.1 Basic principles

The costs for each system configuration are based on component cost data. Costs data have been obtained from literature, vendor quotes and personal communication with experts. For components also present in BIG/CC installations, the FOB (Free On Board) price is multiplied with specific percentages to obtain the installed costs (Faaij, 1997). For components not used in BIG/CC installations, it is assumed as a rule of thumb that the FOB price should be tripled to obtain the installed costs.
The size of each component is based on energy and mass balances as obtained by the Aspen modelling. Scaling factors were used for up- or downscaling. All system components costs are strongly influenced by scale. Economies of scale are calculated per component by using the following relation.
$[Costs]_b/[Costs]_a = ([Size]_b/[Size]_a)^r$, where r is the scaling factor. For biomass systems r is usually between 0.6 and 0.8.

A maximum size for each unit is assumed. Scaling up beyond this maximum size is done by assuming multiple units. The gasifiers used have different maximum sizes.
The maximum size of a gasifier is mainly determined by two factors, whether the gasifier operates at elevated pressure and whether the plant is located near a harbour. If road transport is considered, the dimensions of the road are of importance. When transport of the gasifier to location can be done over water, virtually no limitation is imposed. When road transport is considered, the maximum scale for an atmospheric gasifier is estimated at 100 MWth, while for a pressurised gasifier this 'breakthrough' scale is 250 MWth (Salo 1999) (Pierik 1995). Estimates differ however. Technical director Mark Paisley of BCL estimated a single train BCL gasifier could be used for up to 2000 tons/day (440 MWth). The BCL gasifier has some advantage because it consists of two separate chambers, but it does operate at atmospheric pressure. For coal gasification pressurised circulating fluidised beds have been produced up to 400 MWth (Hogendoorn 2000). Exact maximum scales cannot be given because real world data are not available, but it can be expected that pressurised gasification will have higher maximum scales.

For the TPS, BCL, IGT and EP gasifiers this maximum size is 122, 200, 400 and 400 MWth HHV respectively (see figure 8). Again, it could be that (pressurised) biomass gasification has higher maximum scales but influence on overall economics will be relatively small. The scale investigated in this paper is in the range of 122-1600 MWth HHV.

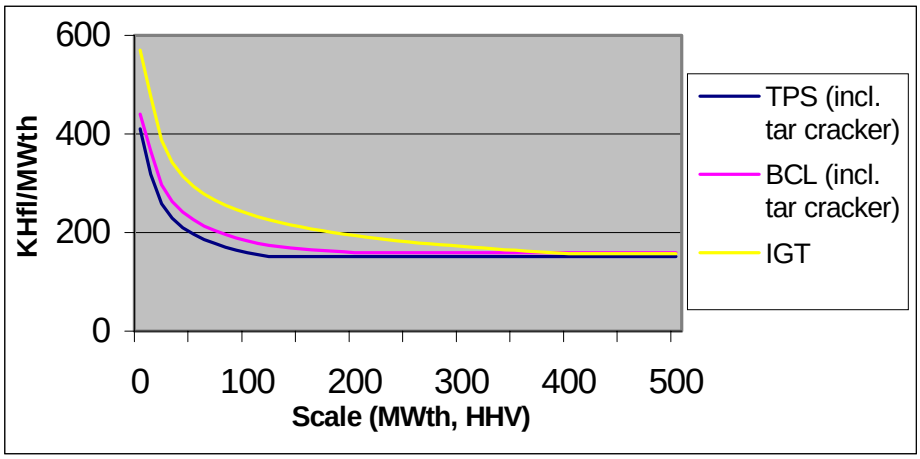

Figure 7: effect of scaling factor and maximum size of gasifiers

4.2 Calculation of production costs of FT liquids

4.2.1 Investment costs

The calculation of the overall total investment costs is done on basis of the cost data as presented in table 12. Table 13 shows the total calculated investment costs for the various concepts.

Table 12: basic costs for all units used with their maximum size (base costs are in relation to base scales). Costs in MHfl (1 Hfl ≈ 2.1 US$)

	Base cost	Scale factor	Base scale	unit	maximum size (for scales considered)
Pre-treatment:					
conveyers (1)	0,7	0,8	69,54	MWth LHV	367
grinding (1)	0,9	0,6	69,54	MWth LHV	367
storage (1)	2,2	0,65	69,54	MWth LHV	367
dryer (1)	16,2	0,8	69,54	MWth LHV	367
iron removal (1)	0,7	0,7	69,54	MWth LHV	367
feeding system (1)	0,8	1	69,54	MWth LHV	367
Gasifiers:					
gasifier TPS (1)	6,8	0,7	69,54	MWth LHV	105
gasifier BCL (incl. feeding) (2)	27,3	0,7	400	MWth HHV	200
gasifier EP (3)					400
gasifier IGT (3)	63	0,7	400	MWth HHV	400
Gas cleaning:					
tar cracker (1)	6,8	0,7	69,54	MWth LHV	105
cyclones (1)	5,4	0,7	69,54	MWth LHV	367
gas cooling (1)	6,2	0,7	69,54	MWth LHV	367
baghouse filter (1)	3,4	0,65	69,54	MWth LHV	367
condensing scrubber (1)	5,4	0,7	69,54	MWth LHV	367
hot gas cleaning (4)	30		400	MWth HHV	none
Compressors:					
compressor (5)	25,2	0,85	13,2	MWe	none
Combined cycle:					
gas turbine (1)	16,1	0,7	25	MWe	none
modifications turbine LCV gas (1)	8%	0,7	69,54	MWth LHV	none
HRSG (1)	7,1	0,8	47,5	tonne/hour	none
Steam turbine + condenser (1)	9,4	0,7	12,3	MWe	none
water+steam system (1)	0,9	0,9	49,5	tonne/hour	none
cooling (1)	2,0	0,3	50,5	tonne/hour	none
Sub1: Total hardware costs					
Instrumentation & control (6)	5% of hardware				
buildings	1,5% of hardware				
grid connections	5% of hardware				
site preparation	0,5% of hardware				
civil works	10% of hardware				
electronics	7% of hardware				
piping	4% of hardware				
Sub2: Investment costs					
engineering (7)	15% of investment costs				
Sub3: Total installed costs(14)					
oxygen plant (8)	48,3	0,75	24	tonne/hour	none
ATR reactor (9)	60,6	0,7			none
Shift reactor (10)	0,95	0,6	2400	kmol/hour	none
Sulfinol D (11)	confidential	0,65			400 MWth input
F-T reactor (12)	35	1	100	MW FT liquid	none
ZnO beds (13)	0,3	1			none

Notes:
(1) Costs figures based on first generation BIG/CC installations, taken from Faaij (1998).
(2) Costs figure taken from (Hamelinck and Faaij 2000).
(3) The Enviro Power gasifier and the IGT gasifier, both operating at elevated pressure, have likewise

designs. Therefore costs are assumed to be the same for both types of gasifiers. Cost figure for IGT taken from (Hamelinck and Faaij 2000).
(4) These costs are an assumption. No real world data are available.
(5) Vendor quote (Sultsor)
(6) Percentages take from Faaij (1998), valid for BIG/CC installations. Adding these percentages to capital costs (free on board, FOB) will result in installed costs.
(7) Engineering is already included in the installed costs for the oxygen plant, ATR, shift, Sulfinol, F-T and ZnO units.
(8) 576 t/d oxygen production of 95 %v purity has a capital cost of 31,000 US$ per t/d, for 1008 t/d the costs are 27,000 per t/d (presentation Air Products, 1991). Oxygen of 99.5% purity requires 5% extra capital costs(presentation Air Products, 1991). Costs are indexed to 1999 US$ using a Consumer Price Index of 0.816 (Sahr, 2000), 1 US$ (1999) is assumed to be 2.1 Hfl.
(9) FOB price for the ATR is 10.1 million US$ (Hamelinck and Faaij 2000). Multiplying with three gives the installed costs.
(10) Calculations were done on basis of NM3 flow and necessary reactor height. Given base unit gives a good representation of this.
(11) Calculations were done on basis of NM3 flow and necessary reactor height.
(12) Main factor used is the amount of CO converted to FT liquids (in MW, HHV based). One catalyst loading is included in these costs. No cost data are available for slurry reactors.
(13) Assuming 1% wt of S (excluding N2) entering the ZnO bed, two guard beds of 3 m3 are necessary. This will take about 2300 kilos of steel. Using a steel price of 20 HFL/kilo, each guardbed will cost Hfl 50,000 (FOB) or 150.000 installed.
(14) For the units below the percentages as discussed in (6) are probably not valid. Therefore installed costs were calculated directly or the FOB price was tripled to obtain the installed costs.

To calculate **overall total investment costs** on top of the installed costs were increased with the following percentages (Faaij, 1998):

building int. 1st year (15)	25%	Installed costs* i.r.
building int. 2nd year	75%	Installed costs* i.r.
project contingency	10%	Installed costs* i.r.
Sub4: Total investment costs		
fees/overheads/profits	10%	of total investment
start-up costs	5%	of total investment
Overall total investment costs		

(15) Interest rate used is 5%

Table 13: overall total investment costs for all concepts, figures in MHfl.

Full conversion concepts	**BCL-R**	**BCLnt**	**BCL**	**IGT-R**	**IGT-hg**	**IGT**	**IGT+**	**EP-R**	**EP-R-nt**	**EP**	**TPS**
Total investment costs	831	763	655	814	753	713	716	944	876	765	810
Idem without Sulfinol D	769	701	614	733	672	640	650	831	763	675	723

Once through concepts, 60% and 80%	**BCL**	**IGT**	**IGT-s**	**IGT+**	**IGT+-s**	**EP**	**TPS**
Total investment costs	589	641	710	624	682	652	695

The IGT concepts have a capital cost breakdown as shown in figure 8. The pre-treatment, gasification with oxygen and gas cleaning account for almost 75% of total capital costs. The use of Sulfinol D will add more than 10% to total capital costs.

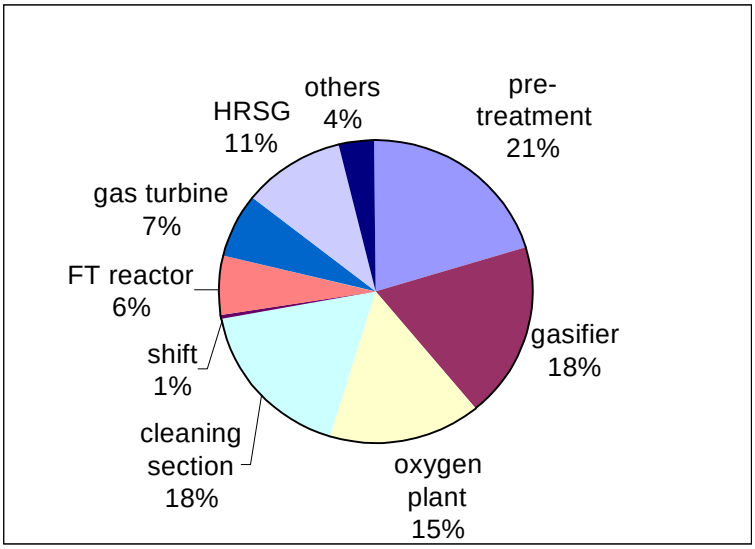

Figure 8: breakdown of installed costs for the IGT 80% conversion concept (400 MWth)

The outcomes of the calculations of energy efficiency and overall total investment costs are used to calculate the production costs of FT liquids, according to the scheme presented in table 16. The breakeven costs are assuming a power price of 0.12HFl/kWh. This includes a premium for green electricity. If electricity has to be bought this can be done at 0.06 Hfl/kWh.

Table 14: outline for calculating production costs per liter FT liquid

Cost factor	unit	Input for 427 MWth
annual depreciation (1)	13.1% of investment	
biomass costs) (2)	4.2 Hfl/GJ (LHV)	10560000
operational costs:		
maintenance	3% of investment	
personnel (3)	1.5 MHFl/100 MWth LHV	367
dolomite (4)	100 Hfl/tonne	25728
waste water treatment (5)	0.45 MHFl/75 MWth LHV	367
NaOH consumption (5)	2.8 KHfl/tonne NaOH	44800
ZnO consumption (6)	70 KHFL/year	
FT cat. Consumption		confidential
insurance	1% of annual depreciation	

Notes:

1. real interest rate = 10%
   depreciation period = 15 years
2. 11,55 GJ/tonne wet, 30% moisture, load factor = 8000 hours/year
3. Faaij (1998), a scaling factor of 0.25 is used.
4. Faaij (1998), dolomite is only used when by a tar cracker is applied.
5. Faaij (1998), waste water treatment and NaOH consumption only if low temperature gas cleaning is applied.
6. Huisman (2000)

4.3 Discussion of results

4.3.1 full conversion concepts

Sulfinol D has a strong impact on overall production costs. For this reason production costs are also calculated without Sulfinol D, assuming the same efficiencies are obtainable. Different will be the LHV of the off gas, which will be lower due the higher CO_2 content. This could endanger the direct use of the off gas in the gas turbine, but co-firing is still possible. Also, overall CO conversion will be lower due to CO_2 build-up in the recycle loop.

Production costs of FT liquids in full conversion concepts, without hydrocracking, are presented in figure 9.

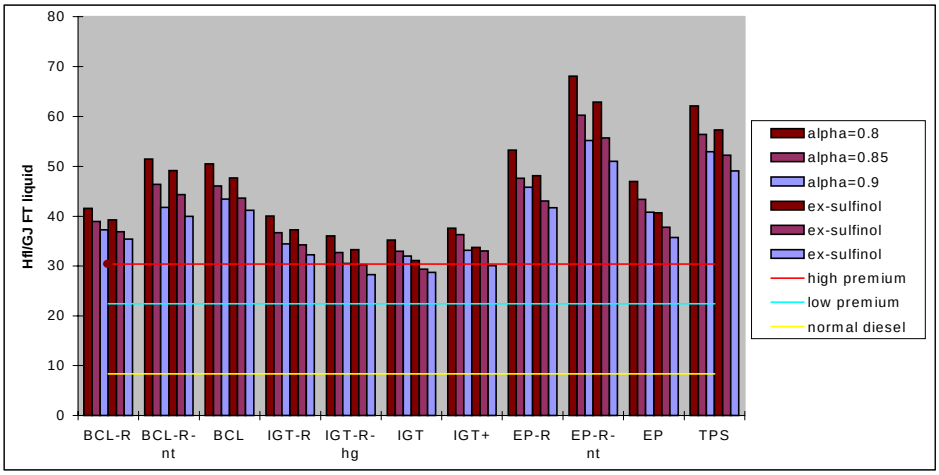

Figure 9: production costs per GJ FT liquid for the full conversion concepts, assuming a power price of 0.12 Hfl/kWh, scale used is 427 MWth HHV.

4.3.2 once through concepts

Production costs for the 60% and 80% conversion once through concepts are shown in figure 10 and 11.

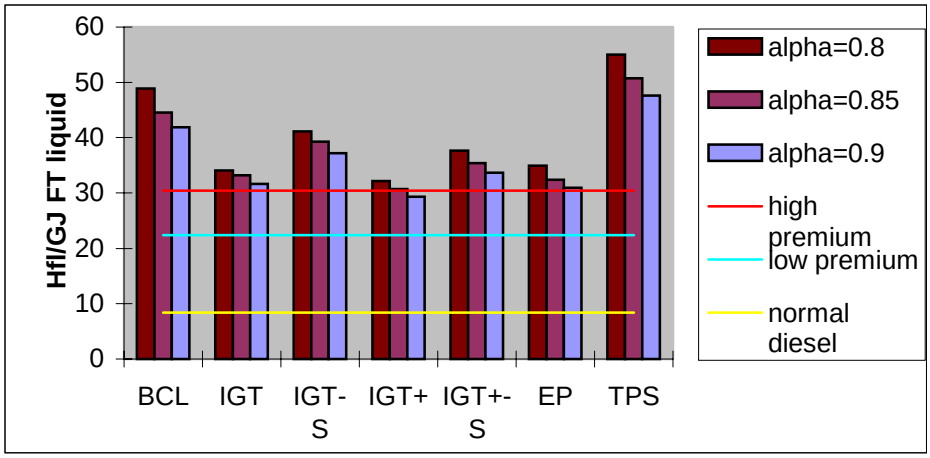

Figure 10: production costs per GJ FT liquid for the 60% conversion once through concepts, assuming a power price of 0.12 Hfl/kWh, scale used is 427 MWth HHV.

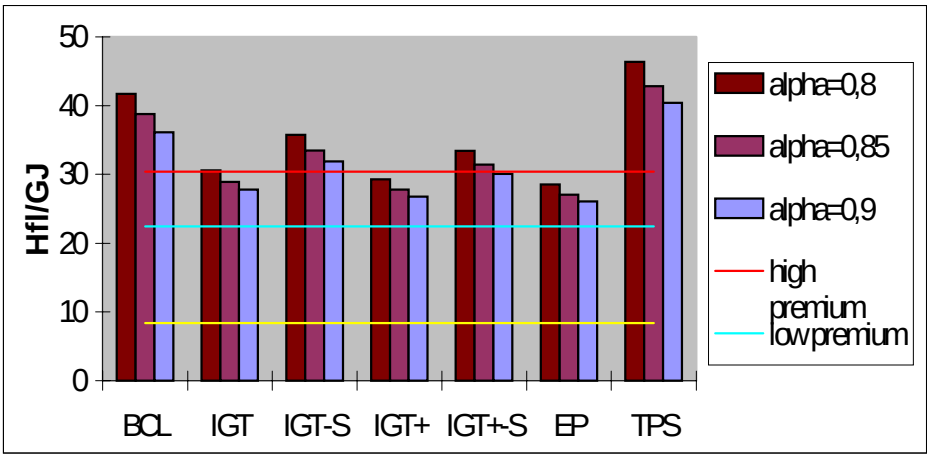

Figure 11: production costs per GJ FT liquid for the 80% conversion once through concepts, assuming a power price of 0.12 Hfl/kWh, scale used is 427 MWth HHV.

4.3.3 Discussion

It is risky to compare the concepts directly. Depending on the amount of inert in the FT reactor, C_{5+} selectivity (α) will differ. For example, the EP once through concept seems slightly better than the IGT once through concept. But IGT, with little inert in its syngas, is more likely to have high α than the EP, having a large amount of inert in its syngas.

One might also conclude falsely that Sulfinol D is a bad option for all full conversion concepts. For instance consider the IGT-R concept. If the reduction of inert by means of Sulfinol D results in α rising from 0.8 to 0.9, it does have a positive effect on production costs. The quantitative impact of the amount of inert on C_{5+} selectivity is not known.

For both groups of concepts, the IGT gasifier, seems to turn out best. Having relatively little inert, the IGT concepts are also very likely to obtain high C_{5+} selectivity. Pressurised systems have a big advantage over atmospheric systems. This is mainly due to the high (energy) costs of compression for concepts using atmospheric gasifiers.

When comparing once through with full conversion one should realise that the once through concepts make use of more advanced FT reactors, with higher once through efficiency. The once through options also make use of a co-fired gas turbine. The co-fired gas turbine has higher energy efficiency and cost-advantages of scale.

4.3.4 hydrocracking costs

The production costs per liter FT liquid are calculated without hydrocracking. A hydrocracking unit is assumed to have installed costs of 17.05 MHfl per 2000 bbl/day or 11352 GJ/C_{5+} per day. Using 13.1% annual depreciation this comes down to 2.23 MHfl/year or 0.54 Hfl/GJ FT liquid. Assuming 98% efficiency for the hydrocracking process and production costs of 30 Hfl/GJ FT liquid, the diesel, kerosene and naphtha would have production costs of 31.16 Hfl/GJ. Also 176 gram H_2/GJ C_{5+} is assumed to be needed, costing 0.35 Hfl/GJ FT liquid[12]. So overall costs for hydrocracking would be Hfl 1.51/GJ on top of Hfl 30/GJ, adding about 5% extra costs.

[12] Assuming H_2 has an energy content of 142 MJ/kg and production costs of 7.17 US$/GJ (based on natural gas, Williams et al. 1995), 176 gram H_2 will have production costs of Hfl 0.35.

4.3.5 competitive production costs

The competitive production costs of 'green' FT liquids strongly depend on the premium for 'green' electricity. In the Netherlands, for green electricity a premium is paid between 0.05 and 0.08 Hfl/kWh (on top of the grid price of 0.06 Hfl/kWh). This corresponds with a premium between 14 and 22 Hfl/GJ. Normal production costs of diesel are 0.30 Hfl/liter or 8.4 Hfl/GJ[13]. Assuming that a similar premium is paid for 'green' FT liquids, 'green' diesel, naphtha and kerosene must have production costs between 22 and 30 Hfl/GJ to be competitive. These values are also shown in figures 9-11.

It can be concluded that pressurised BIG-FT systems have production costs of FT liquids very close to conventional costs plus premium. Atmospheric systems have production costs substantially higher than conventional production costs plus premium. None of the concepts, either atmospheric or pressurised, have production costs competitive with conventional costs at present.

4.4 Sensitivity analysis and cost breakdown

4.4.1 sensitivity analysis of key parameters

The main parameters that affect the final FT fuel costs and their values are shown in table 15; the potential ranges are also shown.

Table 15: main parameters used

parameter	**value**	**range**
Biomass costs	4.2 Hfl/GJ	2.1-8.4
Capital costs	800 million Hfl (varies per concept)	50-175%
Electricity value	0.12 Hfl/kWh	0.06-0.15
Load factor	8000 hours/year	7588-8760
Real interest rate	10 (5% normal interest rate)	6.25-15
Depreciation period	15 years	9.4-22.5

A sensitivity analysis has been performed for these parameters, over the given range. The analysis has been performed for the IGT-R concept at α=0.85, in which case 14% of the total energy output is power. Overall total investment costs for this concept are about 800 MHfl and O&M is about 30 MHfl annually. When power output is high (54% of the total energy output for the EP (full conversion) concept at α=0.85), sensitivity to electricity value is much higher. When all parameters are set at their 'base' value, production costs of FT liquids are 30 Hfl/GJ.

[13] Normal production costs of diesel are (almost linear) dependent on oil prices.

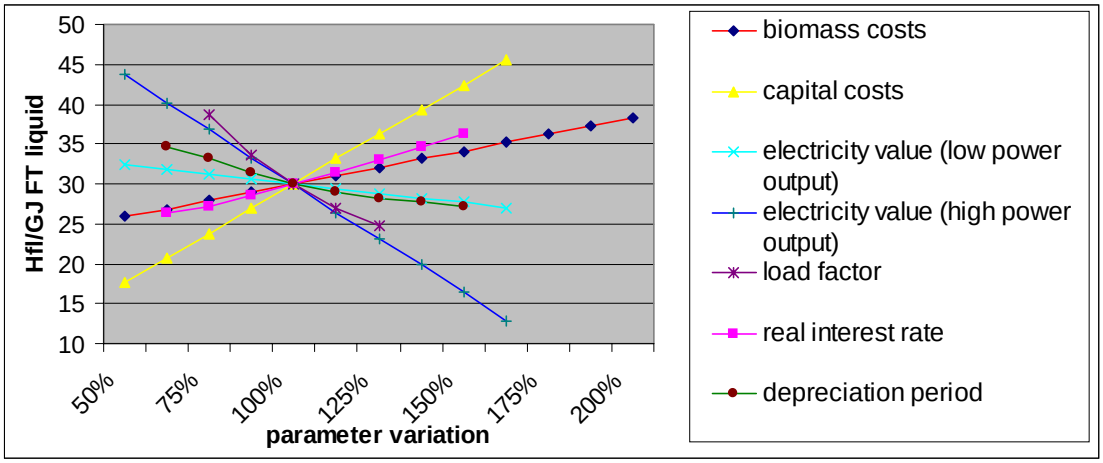

Figure 12: Sensitivity of used parameters to production costs

Scale has a big influence on overall production costs. Overall scaling factors (with respect to overall total investment costs) can be calculated with the scheme presented in table 14, using different maximum size for each unit. When scale is between 100 and 400 MWth the overall scaling factor for the entire factory is approximately 0.74. When scales are larger than 400 MWth, this number is lowered to 0.91. The results for the IGT-R concept is shown in figure 13. It can be concluded that scale effects level off at very large capacities. Scale up can reduce costs of 30 Hfl/GJ at 400 MWth to 26 Hfl/GJ at 1600 MWth, a reduction of 13%. When the capacity is reduced below 400 MWth, costs strongly increase.

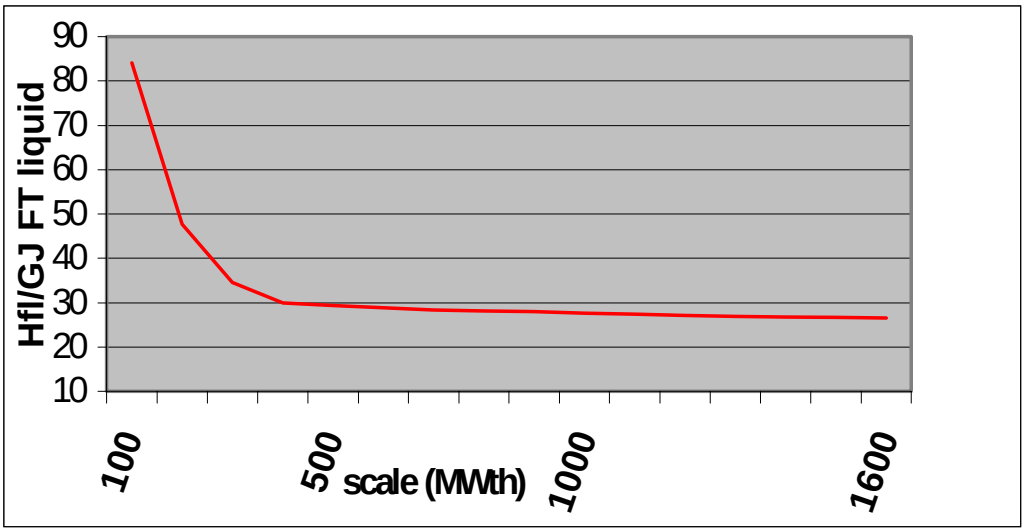

Figure 13: effect of scale on the production costs of FT liquids; production costs of 30 Hfl/GJ assumed at 400 MWth[14]

4.4.2 Short and long term perspective

Results so far deal with technology that could be available on shorter term. However, on the longer term various improvements may be feasible. These include: increasing C_{5+} selectivity, increasing scale, lowering feedstock costs, reduction of investment

[14] Biomass feedstock costs are assumed to be constant; in practise biomass costs could slightly increase for larger scales due to higher logistic costs (Dornburg 1999).

and O&M costs through technological learning and application hot gas cleaning. For the short and long term the following assumptions have been made, covering the range of what is likely to be possible for a first commercial plant and plants built after a period of 30 years.

Table 16: assumptions for the short and long term

Short term (first commercial plant)	Long term (third generation)
• IGT full conversion (40% once through, ex-Sulfinol) is the best concept	• IGT once through 80% conversion (with high efficiency gas turbine) is the best concept[15]
• obtainable α=0.8	• obtainable α=0.9
• scale of the system is 400 MWth	• scale of the system is 1600 MWth[16]
• biomass costs are 4.2 Hfl/GJ	• biomass costs are 4.2 Hfl/GJ[17]
	• technological learning reduces capital costs with 15%[18]

With these assumptions the following cost-breakdown can be given for the short and long term.

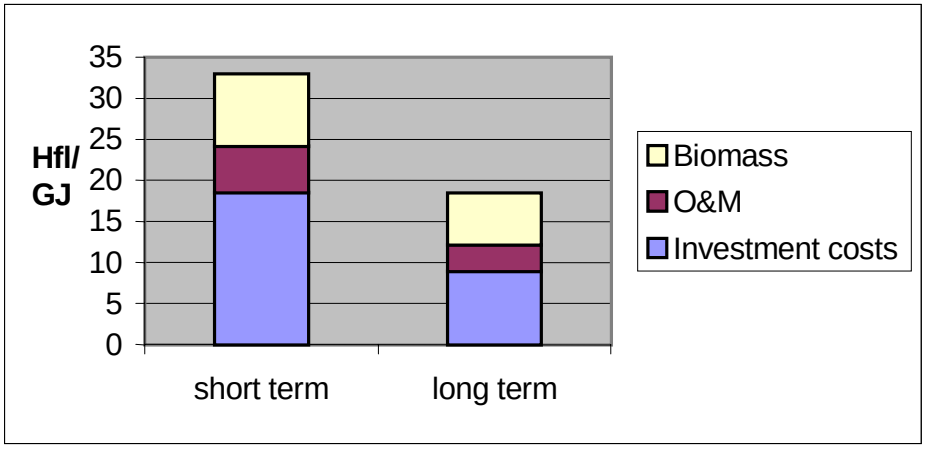

Figure 14: costs breakdown (excluding hydrocracking) for the short and the long term

4.4.3 conclusions

Investment costs have a share of 50% in overall production costs of FT liquids. Reduction of these capital costs for a third generation plant, due to scaling up (12%) and technology learning (15%) therefore have great impact on overall production costs. O&M costs decrease almost proportionally to capital costs. Biomass costs (assumed constant for the short and long term, see footnotes 13 and 16) will decrease per GJ FT liquid due to an increase of overall energy efficiency. Overall energy efficiency will be higher for a third generation plant, due to higher C_{5+} selectivity and higher (once through) CO conversion.

As a result of the reduction in capital costs and biomass costs per GJ FT liquid,

[15] Hot gas cleaning has not been modelled for this concept since only a shift reaction is used for this concept. In that case still some cooling down is needed after the hot gas cleaning and consequently efficiency advantage will be smaller.

[16] An overall scaling factor of 0.91(with respect to overall total investment costs) is used.

[17] In the longer term biomass costs may be lower, but larger scales will increase costs again (see footnote 13).

[18] Technological learning can be assumed on longer term. This can be expressed by a progress curve; such curves are determined by a progress ratio. A progress ratio of x implies that each doubling of cumulative output leads to a (1-x)*100% reduction in costs. A progress ratio of 0.9 is used, applied for a third plant built. This results in 15% lower capital costs (see e.g. Faaij, 1998).

production costs of FT liquids will drop from over 30 Hfl/GJ to below 20 Hfl/GJ. Still, 20 Hfl/GJ is not competitive with conventional costs of diesel, naphtha and kerosene. However, looking at the increase of oil prices over the last years, one may expect the oil price to rise further in the future. This will cause an (almost) proportional increase of conventional production costs of diesel, naphtha and kerosene. So the FT process (either with biomass or natural gas or coal) becomes more attractive with rising oil prices. Rising oil prices will cause production costs of FT liquids with BIG-FT systems to be *closer to*, but not likely *below* conventional costs.

5. Discussion and recommendations

Pressurised systems have much better *overall energy efficiencies* (42-50% LHV) than atmospheric systems (33-40% LHV). This is mainly due too the high electricity consumption of the syngas compressors when atmospheric gasifiers are used.
Both the IGT and EP gasifier come forward as most suitable for BIG-FT systems. High CO conversion, either once through or after recycle, and high C_{5+} selectivity are important for a high overall energy efficiency.

In the *short term*, production costs of FT diesel, naphtha and kerosene will be about 35 Hfl/GJ. Investment costs have a share of 50% in overall production costs of FT liquids. The pre-treatment, gasification (with oxygen) and cold gas cleaning account for almost 75% of total capital costs. Biomass costs are 30% of total production costs, O&M (almost proportionally to investment costs) 20%.

In the *longer term* with large-scale production, high C_{5+} selectivity, high CO conversion and technological learning, production costs of FT liquids could drop to below 20 Hfl/GJ. Reduction of capital costs for a third generation plant, due too scaling up (12%) and technological learning (15%) have great impact on overall production costs. Biomass costs per GJ FT liquid will decrease due to an increase of overall energy efficiency. Overall energy efficiency will be higher for a third generation plant, due to higher C_{5+} selectivity and higher (once through) CO conversion.

When diesel is the desired final product, the FT product requires *hydrocracking*. Besides 60% diesel, 40% naphtha and kerosene are produced. Hydrocracking will add about 5% to production costs. FT liquids are totally free of sulphur, nitrogen, vanadium, asphaltenes and aromatics. FT diesel has a very high cetane number and is of excellent quality. FT naphtha and kerosene have a very low octane number. Finding a suitable outlet for FT naphtha and kerosene might be a matter of concern.

Conventional *production costs* of diesel are about 0.30 Hfl/liter or 8.4 Hfl/GJ (depending very much on oil price). Production costs of ‘green’ FT diesel, naphtha and kerosene (35 Hfl/GJ) are not competitive with conventional prices. In the longer term conventional prices could go up due to higher oil prices, but still ‘green’ FT liquids (19 Hfl/GJ) are not competitive with conventional prices. In the Netherlands, for ‘green’ electricity a premium is paid between 0.05 and 0.08 Hfl/kWh (on top of the grid price). This corresponds with a premium between 14 and 22 Hfl/GJ. Assuming that a similar premium is paid for ‘green’ FT liquids, production costs should be between 22 and 30 Hfl/GJ. In the short term pressurised concepts have production costs close to conventional production costs plus premium. In the longer term however, production costs will be lower than conventional production costs plus premium.
In the Netherlands diesel costs are approximately Hfl 1.80 per liter (≈50 Hfl/GJ) at a gas station. Approximately Hfl 0.90 (≈25 Hfl/GJ) per liter is excise duty, received by the Dutch government and another Hfl 0.27 is VAT (www.shell.com, 1999). If ‘green’, sulphur free FT diesel, naphtha and kerosene would be exempted from excise duty production costs should be below 34 Hfl/GJ. In the short term BIG-FT systems are very close to these production costs, in the long term production costs are likely to be lower.

There are several uncertainties with respect to the *technology status*. A very critical step in the whole system is gas cleaning. It still has to be proven if the gas cleaning section is able to meet the strict cleaning requirements for FT synthesis. Possibly Sulfinol D is required for cleaning purposes, thereby raising production costs. Pressurised (oxygen) gasification systems, having most promising economics and advantages of scale, still need further development. At present, only atmospheric air gasification systems, operating at relatively small scale, have proved to be reliable.

Not all possible *concepts* have been investigated. Separating the C_3-C_4 fraction as liquid product from the FT synthesis could be more advantageous than use of this fraction in a gas turbine. Another good option could be to reform the C_1-C_4 fraction in the recycle loop. Also variable gasification temperatures could make other reforming methods, like partial oxidation and steam reforming, possible due to a different C_{2+} content. For the concepts modelled however, reforming did not result in lower production costs. Lowering the pressure in the FT reactor will cause selectivity to drop and conversion will be lower and equipment will have to be bigger, but on the other hand compression costs will also be lower. Using an iron catalyst could reduce production costs due to an internal shift reaction.

The used *syngas compositions* as produced by the different gasifiers have strong influence on overall results. Most data are taken from literature and based on pilot-scale operating experience. The reliability of these data for large-scale gasifiers is not known.

In the long term the efficiency of the concepts will be higher if high selectivity can be combined with high conversion. This could be realised in either fixed bed or slurry reactors. Costs for slurry reactors, which are not available yet, could be lower than for fixed bed reactors and will definitely have better economies of scale. Heat integration can also be improved. Power generation in the gas turbine will improve if used on large scale. Hot gas cleaning has high potential of improving efficiency, but uncertainty about developments and costs of this promising option are substantial.

Recommendations for further actions and research are:

- The gas cleaning section needs special attention. Proper data sets of contaminants in the syngas must be made, with high detection accuracy. Hot gas cleaning is promising, but will require even more development before sufficient cleaning is guaranteed.
- Pressurised biomass gasification must be developed for large-scale plants, but first demonstration units will have to be built.
- For the use of biomass syngas in the FT synthesis, high liquid selectivity is desirable. The FT process (either a fixed bed or a slurry process) needs to be configured to fulfil this need. Both Shell and Sasol are performing much research on the FT synthesis.
- Development of sustainable forestry is necessary to ensure a large enough supply of clean wood. Efforts must be made to create a working biomass market and reduce prices of biomass, from various sources, over time.
- It should be investigated what the outlets are for FT naphtha and kerosene and if a premium for ‘green’ fuels will be paid for these products.

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Appendix 1: gas cleaning

Method A: 'wet' low-temperature cleaning

(Ree, Oudhuis et al. 1995), (Faaij 1997).

This method consists of several consecutive steps.

-tar cracking in a high-temperature tar cracker (optional) :

At high pressure tars, if formed, are cracked spontaneously, so this application is only necessary when gasification is done at atmospheric pressure. The tars produced in the gasifier are cracked catalytically to smaller compounds. The cracked tars are recycled to the gasifier.

-primary particulates separation in high-temperature devices:

Large particulates are removed (> 5-10 micron). A recycle of the bed material is possible to improve the carbon efficiency. A cyclone separator is most commonly used for primary particle separation, mainly because its capability for high throughput which is necessary when temperature (and pressure) is high. Furthermore its construction is simple and solid.

-cooling and secondary particle removal:

The gas is cooled so that it can pass through a bag filter unit; the filter unit has dust collection efficiencies typically in the 99.7-99.9% range. Operating temperature varies from 120 to 300 °C. The alkalis present in the syngas will condense on the particles when the syngas is cooled below ~450°C.

-syngas scrubber (quench):

Water is used to condense and to dissolve contaminants, mainly NH_3, HCl (and other halogens) and additional hydrocarbons. This depends on the solubility of the contaminant in water. H_2S and CO_2 hardly dissolve in water. Soot and metals are assumed to be washed out completely. NH_3 will be removed to below 1 ppmv. HCN will be removed to below 4 ppmv. All of the HCl will dissolve in the scrubber water with caustic addition (NaOH). The water from the scrubber containing the contaminants (in the form of NaCl at this point) will be transported to a wastewater treatment plant. Assumed is NaOH is recovered from the scrubber as a solid salt. The cold water used to scrub the syngas will cool the syngas. Depending on exit temperature and pressure most of the water vapour in the syngas will condense and exit the gas.

-hydrolisation unit: hydrolysation of COS and HCN

COS will be hydrolysed catalytically (in a separate unit) to H_2S at a desired temperature of 135°C, co-producing CO_2:

$COS + H_2O \leftrightarrow CO_2 + H_2S$

COS concentration will be < 1 ppm afterwards. This is necessary if COS will provoke problems further downstream when H_2S is being processed. HCN will simultaneously be hydrolysed catalytically, to NH_3 or CH_3NH_2.

-Sulfinol D (possibly applied downstream for CO_2 removal)

Sulfinol D or M, licensed by Shell, can be used downstream to remove CO_2. This process also has purifying qualities on some contaminants. All sulphur compounds (H_2S, S, COS, CS_2, mercaptans and organic sulphides and disulphides) are removed to ppm level. Sulfinol M removes 50-80% of the COS in the syngas. If a higher degree of COS removal is necessary, Sulfinol D should be selected. In addition, Sulfinol will remove NH_3, HCN, O_2 and volatile heavy metals such as mercury, selenium and arsenic. (Okimoto and Gadeholt 1998)

So applying Sulfinol has some implications on the upstream cleaning section. The hydrolysation unit will become redundant, since Sulfinol removes COS and HCN. Adding NaOH or H_2SO_4 to the scrubber medium might also become redundant, but this is not quite clear. A ZnO bed is still required for the removal of sulphur to ppb level. One thing to keep in mind is that Sulfinol would be applied prior to the FT section. Applying Sulfinol earlier upstream does not make sense since its main goal is CO_2 removal and syngas processing will induce new CO_2 in the syngas. Cleaning requirements for reactors used for the syngas processing, less severe than FT requirements though, must still be obtained solely by the cleaning section without the aid of Sulfinol.

Method A: results

Particulates:

The cyclone separator and the bag filter unit, with a 99.99% dust collecting efficiency, will remover the bigger part of the particulates. The remainder of the soot will be washed out in the scrubber (quench).

Nitrogen:

Nitrogen, present in the syngas as NH_3 and HCN, can be removed to 1 ppmv and 4 ppmv respectively in the scrubber. HCN will be hydrolysed catalytically to NH_3 or CH_3NH_2 in the hydrolisation unit. Adding an acid to the scrubber water (H_2SO_4) ammonia removal will be more thorough. It has to be investigated further if the FT catalyst requirements are met.

Sulphur:

If a tar cracker is used, a major part of the sulphur (present in the syngas as H_2S and COS) will be absorbed if a dolomite catalyst is used (CaS is formed). COS is hydrolysed catalytically to H_2S in the hydrolisation unit. Adding a basic (NaOH) to the scrubber water will remove sulphur to ppm level (Faaij 1997). This is possible in a two stage scrubber. Problem with biomass derived syngas is the reducing environment, which might complicate sulphur removal. The ppm obtained level isn't satisfactory for FT purposes.

The two options further for H_2S removal are Rectisol (see: acid gas removal) and ZnO guard beds. Selexol (see: acid gas removal) is not favourable for H_2S removal, as is shown in a comparative study done at Shell Research and Technology Centre at Amsterdam (Klinkenbijl, 2000). Since Rectisol is a relatively expensive method, it is only favourable at large scales (Klinkenbijl, 2000), larger than the scales appropriate for biomass based systems. As a result ZnO guard beds have the advantage. ZnO beds with Fe_2O_3 (100-250 °C) have efficiencies >99% and very low outlet sulphur concentrations. The ZnO catalyst is not regenerable (which is normally done with oxygen) and since the ZnO is therefore permanently consumed a reactor configuration with two or three beds in series is used (Klinkenbijl and Loerakker 2000). That way no down time is expected. ZnO beds are unattractive if the H_2S > 20 ppmv in the feed. The scrubber with NaOH will readily meet this in case of using RDF, when poplar wood is used the NaOH might be redundant.

Alkalis (Na, K, Li, V and alkali sulphates):

It is assumed that the gas cleaning system proposed is sufficient to remove alkalis. (Faaij 1997).

Chlorine:
More than 90% of the chlorine, present in the syngas as HCl, will be absorbed by the dolomite if a tar cracker is used (Faaij 1997). HCl will also react with the particles that are later removed by the bag filter. The remaining HCl will dissolve in the scrubber medium.

Heavy metals:
Generally the metals will condense on particulates and be removed by the baghouse filter. The actual behaviour of the metals has not been studied in detail.

Tars
The most common reason for biomass systems to fail is a too high amount of tars in the syngas. This amount is difficult to measure, because it depends on the method used. Tars in the syngas can condense to tar drops, thereby polluting vessels and heat exchangers or they can polymerise and change the properties of the gas (Neeft 2000). The condensing of the tars does not arise if the syngas is not cooled below 500°C, but this has to be done at same stage of the whole biomass-FT process.
In (Katofsky 1993) it is stated that the syngas should not be cooled lower than 400 °C before entering the scrubber. Below this temperature some hydrocarbons will condense on the recovery equipment upstream the quench vessel. However, in (Faaij, Ree et al. 1998) the syngas is cooled to 180 °C for the secondary particle removal. During the cooling down condensable hydrocarbons (and alkali components and other (heavy) metals) are assumed to condense on the surface of the particles. Condensing on the vessel itself is not discussed. This difference might be due to the tar cracker used, but in (Katofsky 1993) a tar cracker is also used for some system configurations. So some indistinctness remains on this subject.
If the syngas is not cooled below 300 °C, a bag filter can't be applied. The type of fibre material and organic coatings used in the filter bag construction imposes this temperature limit. As a result, when the quench is entered at 400 °C, a bag filter may not be useable. An electrostatic precipitator could be used in this case for secondary particle removal (see: hot gas cleaning). This device has a widespread use, efficiencies of 99.85% and is able to operate at temperatures up to 450 °C.

The entire 'wet' low temperature cleaning method is represented schematically in figure 15.

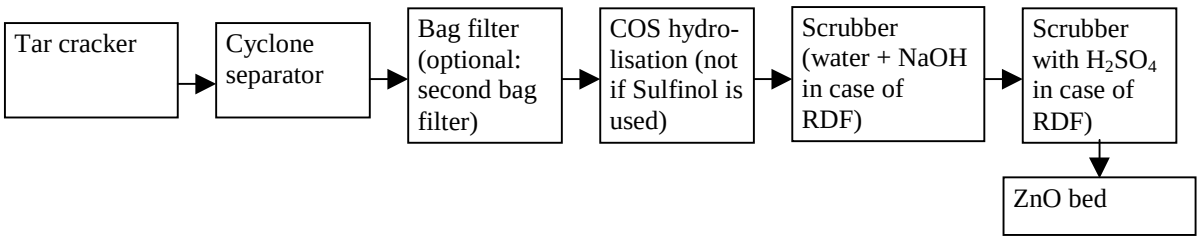

Figure 15: 'wet' low temperature cleaning, schematic view

Method B: 'dry' high-temperature cleaning (hot gas cleaning)
(Ree, Oudhuis et al. 1995) and (Faaij 1997), (Faaij 1999).
This method is attractive because all processes can be operated at high temperature. Therefore the syngas needs not be cooled, which means no heat is lost. But it has to be noted water vapour does not condense and therefore remains in the syngas. The hot gas cleaning method consists of several consecutive steps.

-tar cracking in a high-temperature tar cracker (optional):

This is performed the same way as in method A.

-primary particulates separation in high-temperature devices:

Idem.

-fuel gas cooling and steam production:

The fuel gas is cooled to below 600-700 °C, to condense the alkalis onto the particles in the fuel gas. When ceramic candle filters are used for secondary particle separation, cooling below 850 °C is not desirable.

-secondary particulate separation

Different kind of (advanced) filters can be used, including ceramic candles (up to 99.99% efficiency), electrostatic precipitators (99.85%), high-efficiency cyclone separators (99%), candle filters (99.8%). They operate in the temperature range of 600-850 °C.

-removal other contaminants:

Remaining contaminants require separate cleaning units.

Again, applying Sulfinol downstream will have positive effects on the amount of cleaning needed in the hot gas cleaning section.

Method B: results (most of this information is taken from (Mitchell 1998))

Particulates:

Particulates are removed by the cyclone separator and by the filter device chosen for secondary particle removal. As with 'wet' low temperature cleaning, a second filter may be necessary.

Nitrogen:

NH_3, HCN and tar bound nitrogen are the nitrogen species present in the syngas. NH_3 is the most predominant (>80%). At 800 °C, in the absence of H_2S, 99.5% of the NH_3 in the gas phase was decomposed in an experiment over a nickel catalyst.

Sulphur:

A promising process is under development at Delft University. Using a gamma aluminium-based MnO or FeO sorbent H_2S was purified in an experimental set-up operating at high temperatures to concentrations as low as 1 ppmv. Besides the purification level another key factor for future development is the sorbent capacity. A major problem here is to obtain the strict sulphur purification levels required for FT synthesis. Most systems being evaluated require less strict sulphur cleaning, only having to satisfy emission restraints and therefore developments in thorough sulphur cleaning will probably go at lower rates. Therefore one ore more ZnO guard beds will probably still be necessary, but this will cost no thermodynamical penalty since ZnO beds also operate at high temperature.

Alkali:

When the fuel gas is not cooled below 500-650 °C, alkali compounds are not completely condensed on the particulates and an extra alkali filtering unit is necessary. Using emathlite, the alkalis concentration in the fuel gas can be reduced to 20-50 ppbw. Applying Sulfinol D downstream would also suffice.

Chlorine:

The syngas contains the halogen Cl, and probably F (in case of RDF). Experiments with external fixed bed reactors purified HCl to a value less than 1 ppmv. It can be assumed that in the near future the required level of 10 ppbv will be obtained.

Heavy metals:

At temperatures above 500 °C Pb and some other trace elements are still in the vapour phase. So cooling below this temperature is required (Mitchell 1998). There are no units available today that provide in the required cleaning.

Appendix 2: methane reforming options

The methods described in (Katofsky 1993) are:

-Steam Methane Reforming (SMR):
This conventional method is the most commonly used for producing a syngas from natural gas. Methane conversion is approximately 60-90% for biomass derived syngas, depending on the initial methane concentration and the operating pressure. It is represented by the following highly endothermic reaction:

$CH_4 + H_2O \leftrightarrow CO + 3H_2$ $\Delta H° = +206$ MJ/kmol

Other light hydrocarbons are reformed similarly:
$C_2H_4 + 2H_2O \leftrightarrow 2CO + 4H_2$ $\Delta H° = +210$ MJ/kmol
$C_2H_6 + 2H_2O \leftrightarrow 2CO + 5H_2$ $\Delta H° = +347$ MJ/kmol

Typical operating temperature is 1000-1400 K and pressures vary from 1-3.5 MPa. The composition of the syngas depends on these operating conditions and the steam: carbon ratio. This is partly due to the equilibrium of the water-gas-shift reaction (see below). The H_2/CO ratio to which CH_4 is reformed (3:1) is highest of all reforming options. Heat of the reaction is supplied by burning part of the feedstock.
SMR has various restraints, that cause problems when applied on biomass derived syngas is used (Grootveld 2000). At normal reformer inlet temperature, 500 °C, the CO concentration in the inlet stream cannot exceed 5%, due to the Boudouard reaction. Since biomass-derived syngas has a much higher CO content, temperature has to be raised in order to avoid carbon deposition as a result of the Boudouard reaction. Raising the inlet temperature has a negative side effect. Ethane and higher hydrocarbons, present in the syngas, will be cracked at this high temperatures to methane and carbon. So carbon deposition is not avoided by raising inlet temperature, only the carbon origin is different. As a result, SMR cannot be applied directly after gas cleanup. This problem is also encountered when off gas of a natural gas based FT plant (with likewise syngas composition as biomass derived syngas) is fed to a steam methane reformer.
Various solutions are possible to overcome the problem mentioned. A shift reaction prior to the reformer, reducing CO concentration to less than 5%, will make an inlet temperature of 500 degrees possible. However, this is not practical. When the hydrocarbons are reformed, the H_2/CO ratio will be too high for FT synthesis and a backward shift reaction will be necessary. Another possibility is to apply PSA prior to the reformer. PSA will separate H_2 and CO (and N_2) up to 90%. The other stream will contain the remaining H_2 and CO, but also all CH_4, C_{2+}, CO_2 and (almost) all contaminants. This other stream could well be reformed at an inlet temperature of 500 °C. After the reforming step both streams can be mixed prior to the FT reactor. But the other stream could be used as well to fire a combined cycle, since the caloric value of that stream is relatively high (that way no reforming is necessary). This opposes a new question, but this is beyond the scope of this paper.
SMR has lost its attraction as a result of the above. Other reforming options, may well have the advantage now. A suggestion for further research with respect to steam methane reforming can be made. Raising the gasification temperature will lower the C_{2+} content rapidly above some critical value. Consequently the problems with spontaneous cracking of these C_{2+} could well disappear. But at the same time

gasification efficiency will decrease. In this paper no attention is given to variable gasification temperatures.

-Steam Methane Reforming with CO_2:
The motivation for CO_2 addition is to lower the H_2/CO ratio in the reformat. The reaction is:

$$CH_4 + CO_2 \leftrightarrow 2CO + 2H_2 \qquad \Delta H° = +261 \text{ MJ/kmol}$$

This type of reforming is not applicable for two reasons. First of all a lot of CO_2 (that has to be removed later on) is added to the syngas. Secondly lowering the H_2/CO ratio is not desirable since it already has to be raised for FT synthesis.

- Autothermal Reforming (ATR):
In ATR, only part of the feed is oxidised to produce CO_2 and H_2O; enough to supply the necessary heat to reform the remaining feedstock. It produces a syngas with a lower H_2/CO ratio than conventional steam reforming, depending on steam: carbon ratio and temperature. Operation at higher temperature and pressure than conventional steam reforming is possible. ATR is particularly interesting when oxygen is available, as is the case with direct gasification in an IGT gasifier, but could also be done with air. Reforming with air however will cause extra inert (mainly N_2) to be added to the syngas. A big advantage of ATR is the reduction of capital costs, typically 50-60% less than conventional steam reforming (Katofsky 1993). Operating pressure can be up to 70 bar, outlet temperature vary from 800 to 1,100 °C. The reactor has a high operating flexibility, H_2/CO ratios of the outgoing stream vary between 1.7 and 4.7 when natural gas is reformed.
When a quench is used to clean the syngas the temperature drops to below 100 degrees. Re-heating the gas to such high temperatures will cause a large fraction of CH_4 to be burned in stead of reformed. ATR does have the possibility to shift the H_2/CO ratio by adjusting the steam feed to the reformer.

-Partial Oxidation (POX):
POX is represented by the following reaction.

$$CH_4 + ½O_2 \leftrightarrow CO + 2H_2 \qquad \Delta H° = -35.7 \text{ MJ/kmol}$$

This can either be done with (expensive) oxygen or with (N_2 containing) air. The H_2/CO ratio to which is reformed is close to 2. Partial oxidation with oxygen, like ATR, requires an oxygen plant. Partial oxidation with air again will cause extra inert to be added.
The major problem in applying POX on biomass derived syngas is the high H_2 content. When POX is applied, syngas is mixed with air at 200 degrees C, prior to the reformer. But since the H_2 content is so high, this will lead to spontaneous combustion at the time flows are mixed. As a result, POX is not applicable on biomass derived syngas.

Appendix 3: acid gas removal

CO_2 and H_2S, acid gas contaminants with low solubility in water, can be removed together. Since a scrubber with NaOH and a ZnO bed provide in the required sulphur cleaning an acid gas purification process is not necessary to remove H_2S (Rectisol is too expensive at biomass scales (<1000 MWth) and Selexol isn't suited for H_2S removal, see: gas cleaning). So focus of acid gad removal should be on CO_2 removal, but if the acid gas removal process also removes H_2S, adding NaOH to the scrubber might become redundant. At this point four different acid gas purification processes can be distinguished.

- **Solvent absorption processes**
  Acid gas absorption is a reversible process in which the solvent selectively removes one or more components from the fuel gas in the absorber, after which the components are released from the solvent in the desorber. The 'clean' solvent is then recirculated.

  Physical absorption processes are advantageous when the CO_2 in the syngas has a high partial pressure. Because H_2S and, to a lower extent, COS have a lower saturation vapour pressure than CO_2, physical solvents absorb them preferentially. COS absorption can be promoted by its catalytic hydrolysis to H_2S, as mentioned before. Well-known physical solvent processes are the Rectisol process and the Selexol process. Sasol uses Rectisol for the removal of H_2S and CO_2 in their FT plants (Riesenfeld and Kohl 1974) (Dry 1981). It can move H_2S and COS down to 0.1 ppmv together and CO_2 to 4 ppmv (Processing 1998). The stripped H_2S is sent to a Claus unit (see description below). The advantage of Rectisol is the use of a cheap, stable and easily available solvent (methanol) and a very flexible process with low utilities. The flexibility makes it possible to selectively remove H_2S,without removing CO_2, which is favourable when a Claus unit is integrated in the system. Hydrolysation of COS and HCN is probably not necessary since the Rectisol solvent is capable of removing them to the same extent as H_2S. Questionable is if that will meet the strict FT purification requirements. The process operates at high pressures but at relatively low temperatures.
  Selexol can bring the H_2S concentration down to 1 ppm. CO_2 concentrations of the treated gas are 2-3.5%. Selexol can also be used with an integrated Claus plant.

  Chemical absorption processes can be used for acid gas purification at low to intermediate pressures. Organic chemical solvents used can be amine-based. They have low affinity for the selective absorption of H_2S over CO_2. Another option is the simultaneous absorption of NH_3, H_2S, and CO_2, in an aqueous ammonia and caustic soda solution. This process is capable of removing H_2S and CO_2 down to 0.1 ppmv. A big disadvantage of using inorganic chemical solvents is the possible release of Na, K and V in the product gas, which is not desirable for FT synthesis..
  Combined physical/chemical absorption processes use a mixture of organic solvents. The high capacity for the removal of acid lowers circulation rates necessary and thereby capital and operating costs. A known combined process is Sulfinol. Sulfinol removes H_2S (and all other sulphur compounds) to stringent total sulphur specifications and all or part of the CO_2 (4 ppmv H_2S was readily met in 1990) (Processing 1998). Sulfinol can be used advantageously integrated

with a Claus unit and the SCOT (Shell Claus Off gas Treating) process. SCOT and Claus produce an overall 99.7% recovery of saleable sulphur. As with the Selexol and Rectisol process, Sulfinol combined with a ZnO bed will suffice strict cleaning requirements for H_2S.

Combined physical/chemical absorption processes
These processes use a mixture of organic solvents. These absorbents have high capacities for the removal of acid gases. Because of this high capacity solvent circulation rate is low, resulting in reduced capital and operating costs. Two widely used combined absorption processes are Amisol, developed by Lurgi, and Sulfinol D and M, developed by Shell.

- **Solid phase absorption processes**
  This is the solid phase counterpart to solvent absorption processes. Sorbent capacity for gaseous contaminants is relatively low. This leads to high costs when cleaning of higher concentrations is involved, as is the case for a biomass syngas.

- **Cryogenic separation techniques**
  The low vapour pressures of the acid gas components will lead to their selective separation by gas cooling. No solvent is required, which reduces costs. However, no methods that can obtain the required final H_2S concentration were found.

- **Membrane separation techniques**
  Gas fractionating is achieved by diffusion through a thin membrane barrier separating the feed and the permeate streams. The steady, continuous process simplifies the plant configuration. However, for the selective separation of H_2S from a biomass-derived syngas, no suitable membranes are available yet. Therefore this technology is not an option for the conventional 'wet' low-temperature cleaning.

As a result of the above, of all four acid gas purification processes solvent absorption processes provide the acid gas removal process most suitable to biomass systems.

Two additional processes are distinguishable to further process the waste gas streams from the acid gas treatment processes, to produce elemental sulphur or sulphuric acid. In this way a commercial attractive waste product is produced. As mentioned before, sulphur content in biomass is too low for a Claus unit to be profitable, especially at scales appropriate for biomass systems. To be complete, the Claus process is briefly described below.

- **The conventional Claus process**
  This process recovers H_2S as SO_2 or solid sulphur through oxidation. More than 90% (up to 98%) of the sulphur can be recovered. The tail gas of the Claus unit can be treated in a SCOT unit, where the sulphur is converted to H_2S and recycled back to the Claus unit. When H_2S and CO_2 are removed simultaneously low H_2S concentration will lead to high recovery costs due to a more complex system. Gases with a H_2S content of over 25 % are suitable for the recovery of sulphur in Claus plants (Lurgi). This means that CO_2 recovery will have to be done separately when a Claus unit is used.

Appendix 4: Aspen modelling conditions

The following input data and assumptions need explanation:

- The maximum recycle rate (mol flow recycled gas/mol flow fresh feed) is 2.
- The H_2/CO ratio to which is reformed is 2, the stoichiometric ratio for the FT reaction. The actual H_2/CO ratio used by the FT process is slightly higher (approximately 2.2, depending on α). When a recycle loop is used the effective inlet ratio for the FT reactor will therefore be lower than 2. The FT process operates at 240°C. The inlet stream is preheated to 200°C by the gas leaving the FT reactor. The gas leaving the FT reactor is cooled in steps 50 °C to separate the different carbon fractions. All C_{5+} and H_2O is separated as liquid from the gas. Assumed is no heat is recovered during these cooling down steps, apart from heating the FT inlet stream to 200 °C.
- The FT reactor was assumed to operate at 40 bar. Pressure drop is assumed to be 5 bar. This is more representative for a fixed bed reactor than for a slurry reactor. Impact of this pressure drop is relatively low however since energy demand of the compressor in the recycle loop is small.
- Since the FT reaction is endothermic water is used for cooling down. Steam generated by this cooling down has a pressure of 22 bar (Huisman 2000) and a temperature of 230 °C (10 °C above equilibrium pressure).
- Steam generated at other places in the whole process is heated to 500 °C at 70 bar. If this is not possible, because the quality of the heat is too low, 440 °C at 41.4 bar is used. In the steam turbine this steam is expanded to 22 bar, then the FT steam is added. In steps the gas is expanded further to supply the steam for the ATR (38-44 bar), shift reactor (15.5-44 bar), gasification (1-34 bar, depending on gasification pressure) and biomass drying (12 bar). At a final pressure of 0.04 the steam is at environmental temperature. A maximum of 5 different pressures for the HRSG is used as a practical maximum.
- The gas turbine has an efficiency of 0.91 for the air compressor and 0.89 for the expander, when capacities < 100 MW are considered. This in accordance with a gas turbine type General Electric LM 2500, which can be modified for low caloric value gas. If the LHV of the off gas from the FT process, which is fed to the combustor, is less than 6 MJ/Nm^3 modification of the turbine is necessary. The normal inlet temperature for the expander is 1200 °C, but for low caloric value gas this is set to 1100 °C to assure stable combustion (Ree, Oudhuis et al. 1995). If the LHV of the off gas drops below 6 MJ/NM^3 the pressure of the air is raised from 14 to 16 bar. Assumed is the modified gas turbine can handle all caloric values.
  For the once through concepts with co-firing in a 150 MW gas turbine, efficiencies of 0.91 and 0.9 are used respectively.
- The ‘wet’ cold gas cleaning section is modelled as a single quench with an inlet temperature of 400 °C. This is necessary to avoid condensation prior to the quench (Katofsky 1993). In the quench water of 15 °C is used to cool the syngas to 40 °C below its dew point. According to (Perry 1987) 1-2.7 m^3 water should be used in a scrubber per 1000 m^3 syngas. For the pressurised systems however more water is necessary to obtain the required cooling.
- Heat exchangers are used to heat one flow as another is cooled. Syngas/steam heat exchangers have no practical temperature limits. Syngas/syngas heat exchangers

however cannot heat a stream over 400 °C due too coking problems (Huisman 2000).

All assumptions used for modelling the process in Aspen^{plus} are presented in table 17.

Table 17: input data for Aspen^{plus} modelling. P in bar.

Dryer: per tonne biomass 0.41 tonne steam needed for drying to 10% moisture content, 0.33 for drying to 15%.
'Wet' cold gas cleaning: T_{inlet}=400°C, T_{outlet}=40°C below dew point, Δp=-0.5 at p>30, Δp=-0.3 at 10<p<30, Δp=-0.2 at p<10
Hot gas cleaning: T_{outlet}=450°C, Δp=-1
ATR: p=33-44, Δp=-0.5, T_{inlet}=400-450°C, T_{outlet}=950°C, T_{air}=600°C, T_{oxygen}=300°C
Shift: 15<p<70, Δp=-0.5, T_{inlet}=330°C
FT reactor: p=40, Δp=5, T_{inlet}=200°C, T_{outlet}=240°C, effective H_2/CO_{outlet}>0.4, recycle rate (molflow recycle/molflow fresh) <2, FT steam at p=22 and T=230°C.
Gas turbine: T=1200°C but if LHV< 6 MJ/Nm^3 than T=1100°C, expander: isentropic efficiency=0.89 (0.9 co-fired) mechanical=1, compressor: isentropic efficiency=0.91 mechanical=1, p_{out}=1.2 (0.2 needed for heat exchanger), $T_{flue\ gas}$ (after heat exchanger)> 170°C[1]
Steam turbine: pressure/temperature combinations (T in °C): 70/500, 41.4/440, 22/375 (230°C for FT steam), steam to dryer: p=12, steam to gasifier: p=1-34, steam for shift: p=15.5-43, steam for ATR: p=20-44, outlet pressure of steam turbine=0.04
Heat exchangers: Δp=-0.5 at p>30, Δp==-0.3 at 10<p<30, Δp=-0.2 at p<10, maximum syngas/syngas heating is 400°C[2]

[1] taken from (Faaij 1998). Minimum outlet temperature is due to environmental restraints.
[2]Due to coking problems (Huisman 2000)

Appendix 5: energy balances

Table 18: Overall energy balances of the full conversion concepts for α=0.8

Results full conversion concepts

concept	**BCL-R**	**BCLnt**	**BCL**	**IGT-R**	**IGT-R-hg**	**IGT**	**IGT+**	**EP-R**	**EP-R-nt**	**EP**	**TPS**
alpha=0.8											
overall conversion (%)	82	82	84,4	96	96	87	90	74	74	74	76
inert fresh feed (1-H2+CO)%	35	35	25	1	2	20	21	67	67	66	59
inert in FT reactor (1-H2-CO)%	65	65	55	32	32	52	51	81	81	81	77
Energy efficiency											
input (MW)											
biomass (HHV)	426,7	426,7	426,7	426,7	426,7	426,7	426,7	426,7	426,7	426,7	426,7
biomass (LHV)	367,0	367,0	367,0	367,0	367,0	367,0	367,0	367,0	367,0	367,0	367,0
compressor1	13,5	13,5	13,9	0,0	0,0	0,0	0,0	0,0	0,0	0,0	37,1
compressor2	8,6	8,6	3,9	1,3	1,2	1,3	2,9	8,4	8,4	8,2	10,5
compressor air/oxygen (for ATR)	8,9	8,9	0,0	6,3	6,3	0,0	0,0	18,9	18,9	0,0	0,0
compressor-recycle	2,2	2,2	1,0	0,8	0,8	1,1	1,4	2,6	2,6	2,4	2,5
compressor air/oxygen (gasifier)	0,0	0,0	0,0	8,4	8,4	8,4	8,4	16,6	16,6	16,6	0,0
pumps (kW)	0,6	0,6	0,6	0,6	0,6	0,6	0,6	0,6	0,6	0,6	0,6
output (MW)											
FT liquids (MW, HHV)	133,2	133,2	71,8	139,5	142,5	87,2	84,3	107,5	107,5	69,4	89,3
Gas Turbine	40,6	0,0	56,2	22,9	21,0	60,4	63,3	51,5	0,0	80,8	57,0
HRSG	34,6	20,1	28,1	28,7	32,9	38,5	35,3	48,1	40,4	37,9	31,3
efficiency (%) (HHV)	**40,9**	**28,0**	**32,1**	**40,7**	**42,0**	**40,9**	**39,8**	**37,5**	**23,6**	**37,6**	**29,7**
efficiency (%) (LHV)	**45,1**	**30,1**	**35,9**	**44,7**	**46,1**	**46,0**	**44,6**	**41,6**	**25,4**	**42,4**	**32,9**
net electricity production	**41,4**	**-13,6**	**65,0**	**34,2**	**36,6**	**87,5**	**85,4**	**52,5**	**-6,7**	**90,9**	**37,6**
LHV off gas FT (MJ/Nm3)	*5,9*	*5,9*	33,9	34,2	34	30	27,6	*2,2*	*2,2*	7,5	*5,8*
efficiency gas turbine (%)	36,7		26,1	35,1	32	30,9	32,1	46,5		35,5	33,8
efficiency combined cycle (%)	52,7		37	49,9	43,9	47,3	49,9	56		48	44,1
steam tonnes/hour	194,5		174,3	194,1	203,0	189,0	216,0	242,2	191,7	201,6	175,6

Oxygen of 99.5%v used. This requires 350 kWh/tonne.

16,6 MWe needed for compressing air for the EP gasifier. Based on 'Long term', where 15,17 MW is necessary for a 390 MWth (HHV) system.

Table 19: Overall energy balances of the full conversion concepts for α=0.85

Results full conversion concepts

concept	**BCL-R**	**BCLnt**	**BCL**	**IGT-R**	**IGT**	**IGT+**	**EP-R**	**EP-R-nt**	**EP**	**TPS**
alpha=0.85										
overall conversion (%)	77	77	80	93	83	82	71	71	71	72
inert fresh feed (1-H2+CO)%	35	35	25	1	20	21	67	67	66	59
inert in FT reactor (1-H2-CO)%	61	61	54	32	50	52	81	81	80	76
Energy efficiency										
input (MW)										
biomass (HHV)	426,7	426,7	426,7	426,7	426,7	426,7	426,7	426,7	426,7	426,7
biomass (LHV)	367,0	367,0	367,0	367,0	367,0	367,0	367,0	367,0	367,0	367,0
compressor1	13,5	13,5	13,9	0,0	0,0	0,0	0,0	0,0	0,0	37,1
compressor2	8,6	8,6	3,9	1,3	1,3	2,9	8,4	8,4	8,2	10,5
compressor air/oxygen (for ATR)	8,9	8,9	0,0	6,3	0,0	0,0	18,9	18,9	0,0	0,0
compressor-recycle	2,2	2,2	1,0	0,8	1,1	1,4	2,6	2,6	2,4	2,5
compressor air/oxygen (gasifier)	0,0	0,0	0,0	8,4	8,4	8,4	16,6	16,6	16,6	0,0
pumps (kW)	0,6	0,6	0,6	0,6	0,6	0,6	0,6	0,6	0,6	0,6
output (MW)										
FT liquids (MW, HHV)	149,5	149,5	81,5	161,9	98,7	102,2	121,8	121,8	78,7	101,2
Gas Turbine	35,5	0,0	53,1	15,4	56,7	59,3	46,6	0,0	77,9	53,1
HRSG	31,6	17,5	26,6	26,4	36,9	29,0	51,4	41,5	36,1	30,1
efficiency (%) (HHV)	**42,8**	**31,2**	**33,3**	**43,7**	**42,4**	**41,5**	**40,5**	**27,2**	**38,6**	**31,3**
efficiency (%) (LHV)	**47,0**	**33,5**	**37,1**	**47,7**	**47,4**	**44,9**	**44,8**	**29,4**	**43,4**	**34,5**
net electricity production	**33,3**	**-16,3**	**60,4**	**24,4**	**82,2**	**75,0**	**50,9**	**-5,5**	**86,2**	**32,5**
LHV off gas FT (MJ/Nm3)	*5*	*5*	32,7	31	28,2	26,9	*1,9*	*1,9*	7,2	*5,4*
steam tonnes/hour	189,6		148,5	184,7	180,8	186,6	228,2	196,7	179,1	160,9

Table 20: Overall energy balances of the full conversion concepts for α=0.9

Results full conversion concepts

concept	**BCL-R**	**BCLnt**	**BCL**	**IGT-R**	**IGT**	**IGT+**	**EP-R**	**EP-R-nt**	**EP**	**TPS**
alpha=0.9										
overall conversion (%)	78	78	80	95	82	82	71	71	71	72
inert fresh feed (1-H2+CO)%	35	35	25	1	20	21	67	67	66	59
inert in FT reactor (1-H2-CO)%	61	61	52	30	48	48	81	81	81	76
Energy efficiency										
input (MW)										
biomass (HHV)	426,7	426,7	426,7	426,7	426,7	426,7	426,7	426,7	426,7	426,7
biomass (LHV)	367,0	367,0	367,0	367,0	367,0	367,0	367,0	367,0	367,0	367,0
compressor1	13,5	13,5	13,9	0,0	0,0	0,0	0,0	0,0	0,0	37,1
compressor2	8,6	8,6	3,9	1,3	1,3	2,9	8,4	8,4	8,2	10,5
compressor air/oxygen (for ATR)	8,9	8,9	0,0	6,3	0,0	0,0	18,9	18,9	0,0	0,0
compressor-recycle	2,3	2,3	0,9	0,8	1,1	1,4	3,7	3,7	2,4	2,5
compressor air/oxygen (gasifier)	0,0	0,0	0,0	8,4	8,4	8,4	16,6	16,6	16,6	0,0
pumps	0,6	0,6	0,6	0,6	0,6	0,6	0,6	0,6	0,6	0,6
output (MW)										
FT liquids (MW, HHV)	166,3	166,3	89,7	181,3	109,3	105,2	133,9	133,9	86,5	111,3
Gas Turbine	29,9	0,0	50,5	8,9	53,4	56,7	42,5	0,0	75,4	49,8
HRSG	25,2	19,3	25,3	23,7	33,1	32,4	47,7	41,7	35,4	28,5
efficiency (%) (HHV)	**44,0**	**35,6**	**34,3**	**46,0**	**43,2**	**42,4**	**41,2**	**29,9**	**39,7**	**32,6**
efficiency (%) (LHV)	**48,0**	**38,2**	**38,1**	**50,1**	**48,2**	**47,3**	**45,4**	**32,2**	**44,5**	**35,8**
net electricity production	**21,3**	**-14,6**	**56,5**	**15,2**	**75,1**	**75,7**	**42,0**	**-6,5**	**82,9**	**27,6**
LHV off gas FT (MJ/Nm3)	4,1	4,1	31,6	24,7	26,6	24,9	1,693	1,693	6,964	5,124
steam tonnes/hour	155,9		144,6	169,9	168,5	183,3	244,8	194,8	175,0	156,0

Table 21: Overall energy balances of the once through concepts (60% conversion) for α=0.8

Results once through concepts, 60% conversion

concept	**BCL**	**IGT**	**IGT-s**	**IGT+**	**IGT+-s**	**EP**	**TPS**
alpha=0.8							
overall conversion (%)	60	60	60	60	60	60	60
inert fresh feed (1-H2-CO)%	46	48	20	49	21	74	67
Energy efficiency							
input (MW)							
biomass (HHV)	426,7	426,7	426,7	426,7	426,7	426,7	426,7
biomass (LHV)	367,0	367,0	367,0	367,0	367,0	367,0	367,0
compressor1	13,5	0,0	0,0	0,0	0,0	0,0	37,1
compressor2	5,2	1,6	1,0	4,7	3,0	8,1	12,0
compressor air/oxygen (gasifier)	0,0	8,4	8,4	8,4	8,4	16,6	0,0
pumps	0,8	0,8	0,8	0,8	0,8	0,8	0,8
output (MW)							
FT liquids (MW, HHV)	53,6	63,3	63,3	62,1	62,1	58,8	74,5
Gas Turbine	*150,0*	*150,0*	*150,0*	*150,0*	*150,0*	*150,0*	*150,0*
Gas Turbine: nat.gas	*84,9*	*74,2*	*75,3*	*73,9*	*75,0*	*60,3*	*85,2*
Gas Turbine: off gas	65,1	75,8	74,7	76,1	75,0	89,7	64,8
HRSG	*69,2*	*66,5*	*66,7*	*68,8*	*69,1*	*63,0*	*71,5*
HRSG: gas turbine part	*59,2*	*57,5*	*57,6*	*58,9*	*59,0*	*54,9*	*54,9*
HRSG: biomass part	35,7	38,1	37,7	39,8	39,6	40,9	40,3
efficiency (%) (HHV)	**31,6**	**39,0**	**38,8**	**38,4**	**38,5**	**38,4**	**30,4**
efficiency (%) (LHV)	**35,8**	**44,1**	**43,9**	**43,5**	**43,6**	**43,6**	**33,9**
net electricity production (MW)	**81,3**	**103,0**	**102,2**	**101,9**	**102,4**	**105,1**	**55,2**
LHV off gas FT (MJ/Nm3)	15,8	10,8	22,5	10,6	21,6	5,5	4,5
LHV mixed gas (MJ/Nm3)	20,8	16,0	26,3	15,8	25,7	8,3	8,6
efficiency gas turbine (%)	32,0	36,4	36,1	36,4	36,2	36,8	35,4
efficiency combined cycle (%)	46,8	52,5	52,2	53,1	52,9	52,3	52,3
nat. gas (kmol/hour)	1196	1046	1062	1042	1057	850	1201
steam tonnes/hour	279	280	281	306	307	270	299

Table 22: Overall energy balances of the once through concepts (60% conversion) for α=0.85

Results once through concepts, 60% conversion

concept	**BCL**	**IGT**	**IGT-s**	**IGT+**	**IGT+-s**	**EP**	**TPS**
alpha=0.85							
overall conversion (%)	60	60	60	60	60	60	60
inert fresh feed (1-H2-CO)%	46	48	20	49	21	74	67
Energy efficiency							
input (MW)							
biomass (HHV)	426,7	426,7	426,7	426,7	426,7	426,7	426,7
biomass (LHV)	367,0	367,0	367,0	367,0	367,0	367,0	367,0
compressor1	13,5	0,0	0,0	0,0	0,0	0,0	37,1
compressor2	5,2	1,6	1,0	4,7	3,0	8,1	12,0
compressor air/oxygen (gasifier)	0,0	8,4	8,4	8,4	8,4	16,6	0,0
pumps	0,8	0,8	0,8	0,8	0,8	0,8	0,8
output (MW)							
FT liquids (MW, HHV)	60,8	71,8	71,8	70,3	70,3	66,7	84,3
Gas Turbine	*150,0*	*150,0*	*150,0*	*150,0*	*150,0*	*150,0*	*150,0*
Gas Turbine: nat.gas	*85,8*	*78,9*	*80,0*	*76,6*	*77,7*	*62,7*	*88,5*
Gas Turbine: off gas	64,2	71,1	70,0	73,4	72,3	87,3	61,5
HRSG	*68,8*	*67,4*	*68,0*	*68,7*	*69,4*	*62,9*	*71,4*
HRSG: gas turbine part	*58,8*	*58,4*	*58,9*	*58,8*	*59,3*	*54,8*	*54,8*
HRSG: biomass part	35,2	36,7	36,6	38,6	38,7	40,0	39,1
efficiency (%) (HHV)	**32,9**	**39,5**	**39,4**	**39,5**	**39,6**	**39,5**	**31,7**
efficiency (%) (LHV)	**37,2**	**44,6**	**44,5**	**44,6**	**44,7**	**44,6**	**35,2**
net electricity production (MW)	**79,8**	**97,0**	**96,4**	**98,1**	**98,8**	**101,8**	**50,7**
LHV off gas FT (MJ/Nm3)	15,4	10,4	21,7	10,2	20,9	5,4	4,3
LHV mixed gas (MJ/Nm3)	20,8	16,0	25,9	15,6	25,3	8,3	8,6
efficiency gas turbine (%)	32,0	35,9	35,6	36,4	36,2	36,8	35,4
efficiency combined cycle (%)	46,7	52,0	51,7	53,1	52,9	52,2	52,3
nat. gas (kmol/hour)	1210	1112	1127	1080	1095	884	1247
steam tonnes/hour	279	280	281	306	307	270	299

Table 23: Overall energy balances of the once through concepts (60% conversion) for α=0.9

Results once through concepts, 60% conversion

concept	**BCL**	**IGT**	**IGT-s**	**IGT+**	**IGT+-s**	**EP**	**TPS**
alpha=0.9							
overall conversion (%)	60	60	60	60	60	60	60
inert fresh feed (1-H2-CO)%	46	48	20	49	21	74	67
Energy efficiency							
input (MW)							
biomass (HHV)	426,7	426,7	426,7	426,7	426,7	426,7	426,7
biomass (LHV)	367,0	367,0	367,0	367,0	367,0	367,0	367,0
compressor1	13,5	0,0	0,0	0,0	0,0	0,0	37,1
compressor2	5,2	1,6	1,0	4,7	3,0	8,1	12,0
compressor air/oxygen (gasifier)	0,0	8,4	8,4	8,4	8,4	16,6	0,0
pumps	0,8	0,8	0,8	0,8	0,8	0,8	0,8
output (MW)							
FT liquids (MW, HHV)	66,8	79,0	79,0	77,3	77,3	73,3	92,8
Gas Turbine	*150,0*	*150,0*	*150,0*	*150,0*	*150,0*	*150,0*	*150,0*
Gas Turbine: nat.gas	*87,8*	*81,2*	*82,3*	*78,9*	*80,0*	*65,1*	*91,2*
Gas Turbine: off gas	62,2	68,8	67,7	71,1	70,0	84,9	58,8
HRSG	*68,8*	*67,4*	*67,9*	*68,7*	*69,4*	*62,9*	*71,4*
HRSG: gas turbine part	*58,8*	*58,4*	*58,8*	*58,8*	*59,3*	*54,8*	*54,8*
HRSG: biomass part	34,4	35,7	35,6	37,8	37,8	39,1	38,1
efficiency (%) (HHV)	**33,7**	**40,5**	**40,3**	**40,4**	**40,5**	**40,3**	**32,8**
efficiency (%) (LHV)	**38,0**	**45,6**	**45,4**	**45,5**	**45,7**	**45,5**	**36,3**
net electricity production (MW)	**77,2**	**93,7**	**93,1**	**95,0**	**95,6**	**98,6**	**47,0**
LHV off gas FT (MJ/Nm3)	15,0	10,0	21,0	9,9	20,3	5,2	4,1
LHV mixed gas (MJ/Nm3)	20,6	15,9	25,6	15,5	25,0	8,3	8,5
efficiency gas turbine (%)	32,0	35,9	35,6	36,4	36,2	36,8	35,5
efficiency combined cycle (%)	46,7	52,0	51,7	53,1	52,9	52,2	52,4
nat. gas (kmol/hour)	1237	1145	1160	1112	1127	917	1285
steam tonnes/hour	279	280	281	306	307	270	299

Table 24: Overall energy balances of the once through concepts (80% conversion) for α=0.8

Results once through concepts, 80% conversion

concept	**BCL**	**IGT**	**IGT-s**	**IGT+**	**IGT+-s**	**EP**	**TPS**
alpha=0.8							
overall conversion (%)	80	80	80	80	80	80	80
inert fresh feed (1-H2-CO)%	46	48	20	49	21	74	67
Energy efficiency							
input (MW)							
biomass (HHV)	426,7	426,7	426,7	426,7	426,7	426,7	426,7
biomass (LHV)	367,0	367,0	367,0	367,0	367,0	367,0	367,0
compressor1	13,5	0,0	0,0	0,0	0,0	0,0	37,1
compressor2	5,2	1,6	1,0	4,7	3,0	8,1	12,0
compressor air/oxygen (gasifier)	0,0	8,4	8,4	8,4	8,4	16,6	0,0
pumps	0,8	0,8	0,8	0,8	0,8	0,8	0,8
output (MW)							
FT liquids (MW)	71,5	84,5	84,5	82,8	82,8	78,5	99,3
Gas Turbine	150,0	150,0	150,0	150,0	150,0	150,0	150,0
Gas Turbine: nat.gas	93,4	83,9	84,8	83,5	84,6	69,5	96,8
Gas Turbine: off gas	56,6	66,1	65,2	66,5	65,4	80,5	53,2
HRSG	71,7	68,3	68,5	70,6	70,9	70,8	73,7
HRSG: gas turbine part	59,3	57,5	57,5	58,9	59,1	55,0	55,0
HRSG: biomass part	34,7	36,2	36,0	37,8	37,6	45,3	38,2
efficiency (%) (HHV)	**33,6**	**41,2**	**41,1**	**40,6**	**40,7**	**41,9**	**33,0**
efficiency (%) (LHV)	**37,7**	**46,4**	**46,2**	**45,6**	**45,7**	**47,2**	**36,5**
net electricity production (MW)	**71,9**	**91,5**	**91,0**	**90,4**	**90,8**	**100,4**	**41,5**
LHV off gas FT (MJ/Nm3)	17,1	11,1	28,7	10,8	27,0	5,3	4,1
LHV mixed gas (MJ/Nm3)	22,6	17,4	30,2	17,1	29,4	8,7	9,1
efficiency gas turbine (%)	32,0	36,4	36,1	36,4	36,1	36,8	35,4
efficiency combined cycle (%)	47,3	53,0	52,6	53,5	53,2	54,2	52,8
nat. gas (kmol/hour)	1316	1182	1195	1177	1192	979	1364
steam tonnes/hour	283	288	288	314	314	279	310

Table 25: Overall energy balances of the once through concepts (80% conversion) for α=0.85

Results once through concepts, 80% conversion

concept	**BCL**	**IGT**	**IGT-s**	**IGT+**	**IGT+-s**	**EP**	**TPS**
alpha=0.85							
overall conversion (%)	80	80	80	80	80	80	80
inert fresh feed (1-H2-CO)%	46	48	20	49	21	74	67
Energy efficiency							
input (MW)							
biomass (HHV)	426,7	426,7	426,7	426,7	426,7	426,7	426,7
biomass (LHV)	367,0	367,0	367,0	367,0	367,0	367,0	367,0
compressor1	13,5	0,0	0,0	0,0	0,0	0,0	37,1
compressor2	5,2	1,6	1,0	4,7	3,0	8,1	12,0
compressor air/oxygen (gasifier)	0,0	8,4	8,4	8,4	8,4	16,6	0,0
pumps	0,8	0,8	0,8	0,8	0,8	0,8	0,8
output (MW)							
FT liquids (MW)	81,0	95,7	95,7	93,8	93,8	88,9	112,5
Gas Turbine	150,0	150,0	150,0	150,0	150,0	150,0	150,0
Gas Turbine: nat.gas	96,5	87,6	88,4	87,1	88,2	72,7	101,0
Gas Turbine: off gas	53,5	62,4	61,6	62,9	61,8	77,3	49,0
HRSG	71,5	68,2	68,4	70,5	70,9	70,5	73,6
HRSG: gas turbine part	59,2	57,5	57,5	58,9	59,1	55,0	55,0
HRSG: biomass part	33,4	34,6	34,5	36,3	36,2	43,9	36,6
efficiency (%) (HHV)	**34,8**	**42,7**	**42,6**	**42,0**	**42,1**	**43,3**	**34,7**
efficiency (%) (LHV)	**38,9**	**47,8**	**47,7**	**47,0**	**47,1**	**48,6**	**38,2**
net electricity production (MW)	**67,4**	**86,3**	**85,9**	**85,3**	**85,8**	**95,7**	**35,6**
LHV off gas FT (MJ/Nm3)	16,4	10,4	27,2	10,2	25,6	5,1	3,8
LHV mixed gas (MJ/Nm3)	22,3	17,2	29,6	16,9	28,8	8,6	9,0
efficiency gas turbine (%)	32,0	36,4	36,2	36,4	36,1	36,8	35,4
efficiency combined cycle (%)	47,3	52,9	52,7	53,5	53,2	54,1	52,8
nat. gas (kmol/hour)	1360	1234	1246	1228	1243	1024	1424
steam tonnes/hour	283	288	288	314	314	278	309

Table 26: Overall energy balances of the once through concepts (80% conversion) for α=0.9

Results once through concepts, 80% conversion

concept	**BCL**	**IGT**	**IGT-s**	**IGT+**	**IGT+-s**	**EP**	**TPS**
alpha=0.9							
overall conversion (%)	80	80	80	80	80	80	80
inert fresh feed (1-H2-CO)%	46	48	20	49	21	74	67
Energy efficiency							
input (MW)							
biomass (HHV)	426,7	426,7	426,7	426,7	426,7	426,7	426,7
biomass (LHV)	367,0	367,0	367,0	367,0	367,0	367,0	367,0
compressor1	13,5	0,0	0,0	0,0	0,0	0,0	37,1
compressor2	5,2	1,6	1,0	4,7	3,0	8,1	12,0
compressor air/oxygen (gasifier)	0,0	8,4	8,4	8,4	8,4	16,6	0,0
pumps	0,8	0,8	0,8	0,8	0,8	0,8	0,8
output (MW)							
FT liquids (MW, HHV)	89,1	105,3	105,3	103,1	103,1	97,8	123,7
Gas Turbine	150,0	150,0	150,0	150,0	150,0	150,0	150,0
Gas Turbine: nat.gas	97,6	90,7	91,5	90,1	91,2	75,6	104,7
Gas Turbine: off gas	52,4	59,3	58,5	59,9	58,8	74,4	45,3
HRSG	70,9	68,1	68,4	70,4	70,8	70,4	73,4
HRSG: gas turbine part	59,2	57,5	57,5	58,9	59,1	55,0	55,0
HRSG: biomass part	32,4	33,3	33,3	35,0	34,9	42,7	35,0
efficiency (%) (HHV)	**36,2**	**43,9**	**43,8**	**43,2**	**43,2**	**44,4**	**36,1**
efficiency (%) (LHV)	**40,4**	**49,0**	**48,9**	**48,2**	**48,3**	**49,8**	**39,7**
net electricity production (MW)	**65,3**	**81,9**	**81,6**	**81,0**	**81,4**	**91,6**	**30,5**
LHV off gas FT (MJ/Nm3)	15,9	9,9	25,8	9,7	24,4	4,9	3,5
LHV mixed gas (MJ/Nm3)	22,0	17,0	29,1	16,7	28,3	8,6	9,0
efficiency gas turbine (%)	32,0	36,4	36,2	36,4	36,1	36,8	35,4
efficiency combined cycle (%)	47,1	52,9	52,7	53,5	53,1	54,1	52,7
nat. gas (kmol/hour)	1375	1278	1290	1270	1286	1065	1475
steam tonnes/hour	286	288	288	314	314	278	309