Production of FT transportation fuels from biomass; technical options, process analysis and optimisation, and development potential

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Colofon

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Executive summary

Fischer Tropsch (FT) diesel derived from biomass via gasification is an attractive clean and carbon neutral transportation fuel, and directly usable in the present transport sector. Previous research indicated that FT diesel could be produced at 9.1 – 16.7 $€_{2002}$/GJ (long – short term). Thus, without exemption of excise duty and vat (11.6 and 3.5 €/GJ in the Netherlands), it cannot compete with fossil diesel (Rotterdam product price over past five years was 2.6 – 7.0 €/GJ, BP 2002). These studies were still of a strong explorative nature, bearing rough assumptions on process equipment performance, ignoring the influence of process conditions on conversion and selectivity in the FT reactor, and rough in the economic analysis. Furthermore these studies listed improvement options to bring the production costs of bio FT diesel to a competitive level. Various essential system components are not commercially available and require development over time.

The underlying study incorporates these improvements and refines the technical and economical analysis. The main system efficiency improvements reside in combined production of FT liquids with power, application of pressurised gasifiers, and in combining high selectivity with high conversion in the FT reactor. Since about 75 % of the investment costs are in the pre-treatment, gasification and gas cleaning section, the influence of the gasification pressure and medium (air / oxygen enriched air / pure oxygen) on economies of both gasifier and downstream equipment are studied more closely. The scale of the conversion system is expected to be an important factor in the overall economic performance, but at the same time the perspectives for large throughput biomass gasifiers are (heavily) disputed. Therefore, special attention is given to the technical perspectives for large-scale pressurised biomass gasification. Furthermore insight is gained in the sensitivity of Fischer-Tropsch catalysts to contamination in the biomass-derived synthesis gas, in necessary gas cleaning equipment to avoid catalyst poisoning and ageing, and in reactor design.

System components necessary for FT production from biomass are analysed and combined to a limited set of promising conversion concepts. The main variations are in gasification pressure, the oxygen or air medium, and in optimisation towards liquid fuels only, or towards the product mix of liquid fuels and electricity. The technical and economic performance is analysed. For this purpose a dynamic model was built in Aspen Plus, allowing for direct evaluation of the influence of each parameter or device, on investment costs, FT and electricity efficiency and resulting FT diesel costs.

A 400 MW_{th} input plant (or 168 $MW_{FT,HHV}$) applying conventional technologies is taken as the *key concept* for further analysis. It consists of a 25 bar oxygen fired gasifier, followed by a tar cracker and wet gas cleaning and a solid bed FT reactor with 70 % once through conversion. The Total Capital Investment or TCI for this concept is about 286 M€, this complies with literature values found for grass roots natural gas fed FT plants. A high conversion (90% of $CO+H_2$) can be realised in once through mode by application of a large reactor. Of course this results in higher capital costs, but since the capital costs of the FT reactor are not crucial for the total capital investment, in total a higher conversion leads to lower FT diesel production costs. Overall efficiencies for the best performing systems are 40 – 45 % on HHV basis. With such systems FT liquids can be produced at 16 €/GJ.

System variations to the above named *key concept* give the following insights:

- Removing the 30 volume % CO_2 fraction prior to the FT reactor improves both selectivity and efficiency, but due to the accompanying increase in investment this does not result in lower product costs.
- The efficiency for the dry gas cleaned concepts is slightly higher than wet cleaned concepts, because the ceramic membrane more effectively shifts (since product is taken away) than the traditional shift reactor, hence less steam is needed. This is outweighed by a slightly higher capital investment, such that the resulting fuel price is the same.
- The oil scrubber effectively returns BTX and tars to the gasifier where they are cracked, but much energy remains in methane, ethane and ethane, resulting in lower FT yields. The performance of concepts with oil scrubber improves when adding a reformer, however, this turns out to be a more expensive combination than a single tar cracker.
- Application of a recycle loop instead of once through does not improve the production costs: The continuous temperature leaps and product leaps (syngas partly to C_{1-4} product, and than back to syngas) lead to both a low selectivity and a low overall efficiency. Furthermore, the capital costs are high.

The FT liquids production costs for the short term are still about two to four times the production costs for Fossil diesel. Fossil diesel costs strongly depend on the oil price, and could go up. Green diesel could be exempted from excise duty and VAT (11.6 €/GJ and 3.5 €/GJ in the Netherlands) to value the environmental benefits of green FT diesel. The combined effect possibly makes FT diesel from biomass competitive with fossil diesel.

On the longer term (~15 years), more cost improvements are foreseen. The combined effects of larger production scale (2000 MW_{th}) and cheaper biomass (2 €/GJ) may bring the FT fuel production costs to 11.5 €/GJ. The technical and economical performance may be improved by for example developments in oxygen production and gasification, application of catalytic tar cracking instead of thermal, increased selectivity towards the desired product in FT synthesis. These technical developments combined with technological learning (capital costs reduce with 15 % for the third generation plant) bring the FT diesel costs to 9 €/GJ. Finally, green FT diesel seems more expensive than other biofuels such as methanol or hydrogen, but preserves its advantages of better applicability in the existing infrastructure.

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1 Introduction

The transportation sector represents about 86 EJ/yr or 21 % of the world's primary energy consumption and this share still increases (EIA 2002). Transportation fuels are, at present, predominately derived from mineral oil, leading to greenhouse gas emissions. Concurrently the geographical imbalance between major producers and large consumers raises questions on energy security and global social stability.

Biomass derived transportation fuels have very low lifecycle CO_2 emissions if the biomass feedstock is sustainably grown (Williams et al. 1995), and partly reduce local air pollution. Fischer-Tropsch (FT) diesel from biomass via gasification is getting growing attention in recent years, as it offers a clean and carbon neutral transportation fuel, directly usable in the present day transportation sector. Furthermore FT diesel is – via onboard reforming – very suitable for fuel cell vehicles, because it is free of fuel cell catalyst poisons.

Previous research (esp. Tijmensen et al. 2002; further den Uil et al. 2001; Arthur D. Little 1999) indicated that FT diesel could be produced at 9.1 – 16.7 €[1]/GJ_{HHV} (long – short term). Thus without exemption of excise duty and vat (11.6 and 3.5 €/GJ respectively in the Netherlands), it cannot compete with fossil diesel (Rotterdam product price over past five years was 2.6 – 7.0 €/GJ, BP 2002). The study of Tijmensen was – though more detailed than previous studies – still of a strong explorative nature, bearing rough assumptions on process equipment performance, ignoring the influence of process conditions on conversion and selectivity in the FT reactor, and rough in the economic analysis. Also, Tijmensen listed improvement options to bring the production costs of bio FT diesel to a competitive level. Various essential system components are not commercially available and require development over time. The present study aims to incorporate these improvements and refines the technical and economical analysis. The main system efficiency improvements reside in combined production of FT liquids with power, application of pressurised gasifiers, and in combining high selectivity with high conversion in the FT reactor. Since about 75 % of the investment costs are in the pre-treatment, gasification and gas cleaning section, the influence of the gasification pressure and medium (air / oxygen enriched air / pure oxygen) on economies of both gasifier and downstream equipment will be studied more closely. The scale of the conversion system is expected to be an important factor in the overall economic performance (see Larson and Marrison 1997; Faaij et al. 1998; Hamelinck et al. 2001; Tijmensen et al. 2002; Sie and Krishna 1999), but at the same time the perspectives for large throughput biomass gasifiers are heavily disputed. Therefore, special attention will be given to the technical perspectives for large-scale pressurised biomass gasification. Previous research also indicated that insight is needed in the sensibility of Fischer-Tropsch catalysts to contamination in the biomass-derived synthesis gas, in necessary gas cleaning equipment to avoid catalyst poisoning and ageing, and in reactor design (Wilhelm et al. 2001), however, many of the ruling uncertainties have been taken away by recent ECN research (Boerrigter et al. 2002).

This study tries to answer *what are/will be the technical and economic performances for bio FT diesel production?*, *what price is feasible in the short, mid and long term?*, *and with which process configurations?* Furthermore, the study will indicate *the RD&D priorities for eventual realisation of the optimal configuration(s)*, and identify *the most probable and most likely successful development paths to get there*.

To address these questions, the state of the art technology is extensively surveyed: In Chapter 2 all necessary system elements are described and their development stage and improvement options indicated. The most promising concepts are presented in a shortlist in Chapter 3. Chapter 4 describes the technical modelling in Aspen Plus, to analyse the energetic performance. One extensive adjustable computer model is constructed, with in its center an FT reactor which incorporates sensibility of conversion extent and liquid selectivity towards different reaction mixtures and reactor conditions. Special attention is given to the heat integration within the proposed plants. Results on efficiencies and stream sizes will be used for economic evaluations in order to calculated fuel prices by factored estimation in Chapter 5. This report is concluded by a discussion and conclusion.

[1] All costs in this report are in $€_{2002}$. Energy is expressed on higher heating value basis.

2 System components

The basic steps in production of FT liquids (den Uil et al. 2001; Geerlings et al. 1999; Tijmensen et al. 2002) are depicted in Figure 2-1. After pre-treatment, biomass is gasified to produce synthesis gas (biosyngas). This gas is cleaned and the composition is modified to fit the specifications for Fischer-Tropsch synthesis in the FT reactor. FT offgas is recycled or combusted to produce electricity. The liquid FT products are treated to yield gasoline (C_5-C_{10} hydrocarbons) and/or diesel.

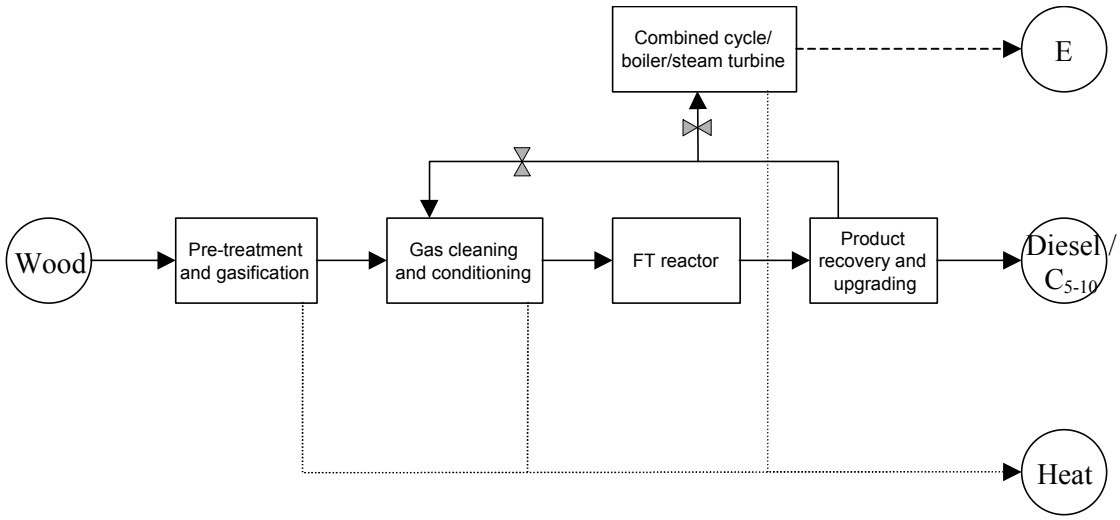

Figure 2-1. Principle of FT liquids producing plant.

In this chapter the separate system units are described following their position in the plant. Improvement options for efficiency or economics, at unit or plant level are indicated.

2.1 Pre-treatment and gasification

Several pre-treatment units precede gasification, such as drying and size reduction. For drying a significant amount of energy is needed, which can be the FT process heat, or extracted from the plant off gas or from the steam cycle. The efficiency of gasification increases with drier biomass (den Uil et al. 2001; Hamelinck et al. 2001; Tijmensen et al. 2002; Faaij et al. 1998; van Ree et al. 1995), however the hydrogen content of the produced syngas decreases, which may be unfavourable for the Fischer Tropsch synthesis downstream, and costs increase quickly with very dry biomass. The resulting effect of a certain extent of drying on the overall plant efficiency/costs is a priori unknown. For BIG CC applications 10 – 15 % is the optimum (Faaij et al. 1998) and in this study a biomass moisture content of 15% is assumed.

To prevent self-ignition and dust explosion of the chipped biomass prior to entering the gasifier the biomass should be made inert. This may effectively be done by addition of steam (only in atmospheric gasifiers) carbon dioxide, or nitrogen.

Many gasification methods are available for syngas production. Based on throughput, cost, complexity, and efficiency issues, circulated fluidised bed gasifiers are very suitable for large-scale gas production. A previous study ruled out direct gasification with air, because the nitrogen dilution strongly increases downstream equipment size (Hamelinck et al. 2001), and it analysed the IGT (Institute of Gas Technology) pressurised direct oxygen fired gasifier, and the BCL (Battelle Columbus) atmospheric indirect fired gasifier. The present study focuses on direct-fired gasifiers and attempts to bring the influence of gasification medium and pressure to light.

The economies of scale are very important for the technologies considered (Faaij et al. 1998). Atmospheric CFB gasification up to 300 MW_{th} is technically feasible, but the upper technical limit to capacity has never been explored as generally fuel supply is considered as limiting factor (Greil 2002). Pressurised gasification may have economic advantages: the gasifier may be much smaller per throughput, so that a larger maximum capacity is expected to be possible. Moreover, as no additional compression is required, the temperature can be kept high for high temperature operations downstream.

At pressures above 25-30 bar gasifiers need to be constructed more heavily (or the throughput should be kept small) and feeding becomes expensive. The present study limits gasification to 25 bar.

Gasification with oxygen as oxidative medium, offers benefits in downstream equipment size, compression energy, and higher partial pressures for relevant components in FT. Pure oxygen is expensive, but oxygen enriched air may combine the advantages of equipment size reduction and a an less expensive medium. In the present study, the extremes 'pure' air and pure O_2 are to be studied, as well as one intermediate. Steam (or CO_2) replaces the heat capacity of nitrogen and serves fluidisation.

2.2 Gas cleaning and conditioning

The gas produced by gasification contains impurities, typical are the organic impurities tars and BTX (benzene, toluene, and xylenes), the inorganic impurities NH_3, HCN, H_2S, COS, and HCl, and furthermore, volatile metals, dust, and soot. Organic contaminations can be removed either by cracking or scrubbing. When the tars and BTX are removed, the other impurities are removed by standard wet gas cleaning technologies or advanced dry gas cleaning technologies.

2.2.1 Tars + BTX removal

In gasification, larger hydrocarbons are formed, generally categorised as 'tars'. When condensing, they foul downstream equipment, coat surfaces and enter pores in filters and sorbents. To avoid this, their concentration must be below dew point at FT pressure. On the other hand they contain a lot of potential CO and H_2. They should thus preferably be cracked into smaller hydrocarbons. Fluidised beds produce tar at about 10 g/m_{NTP}^3 or 1 – 5 wt% of the biomass feed (Boerrigter et al. 2002; Milne et al. 1998). BTX (benzene toluene xylene; or volatile aromatics), accounting for 0.5 volume % of the syngas, have to be removed prior to the active carbon filters, which otherwise sorb the BTX completely and quickly get filled up (Boerrigter et al. 2002).

Three methods may be considered for tar removal/cracking: Thermal cracking, catalytic cracking, and scrubbing. At temperatures above 1000 °C – 1200 °C, tars are destroyed without a catalyst, usually by addition of steam and oxygen. Drawbacks are the need for expensive materials, soot production (Milne et al. 1998) and the low thermal efficiency.

Catalytic cracking (dolomite or Ni based) is best applied in a secondary bed, and avoids the mentioned problems of thermal cracking. However the technology is not yet fully proven (Milne et al. 1998) and the catalyst consumption and costs are matters of concern.

Tars can also be removed at low temperature by advanced scrubbing with an oil based medium (Boerrigter et al. 2002). The tar is subsequently stripped from the oil and reburned in the gasifier. At atmospheric pressures BTX are only partially removed, at elevated pressure (from about 6 bar) BTX are fully removed. The gas enters the scrubber at about 400 °C and leaves the unit at 100 – 80 °C, this allows high temperature heat exchange before the scrubber.

2.2.2 Other contaminants removal

Now tars are absent, other impurities, such as NH_3, H_2S, COS, HCl, volatile metals, dust, and soot, can be removed by standard wet gas cleaning technologies or advanced dry gas cleaning technologies. Maximal acceptable contaminant concentrations for the cobalt FT catalyst are summarised in Table 2-1, together with the effectiveness of two gas cleaning methods.

Table 2-1. Contaminant specification for cobalt FT synthesis[1)], and cleaning effectiveness of wet and dry gas cleaning.

Contaminant	Gas phase Specification	Treatment method & remarks Existing technologies	Dry gas cleaning[5)]
soot (dust, char, ash)	-[2)] 0 ppb[3)]	Cyclones, metal filters, moving beds, candle filters, bag filters, special soot scrubber[3)] Specifications are met.	
alkaline (halide) metals	-[2)] < 10 ppb[3)]	Active coal bed meets specification[2)]	Sorbents Under development
Tar	below dew point at FT pressure[2),] catalyst poisoning compounds < 1 ppmV[2)]	Thermal tar cracker[2)], Oil scrubber[2,6)], Specifications are met[2)].	Catalytic tar cracker, other catalytic operations. Under development.
BTX	below dew point at FT pressure[2)]		
Halide compounds			
HCl (HBr, HF)	< 10 ppb[2,3)]	removed by aqueous scrubber[2,3)] active coal bed meets specification[2)] absorbed by dolomite in tar cracker (if applicable),	In-bed sorbents or in-stream sorbents < 1 ppm Guardbeds necessary
Nitrogen compounds	total nitrogen <1 ppmV[2)] < 20 ppb total N[3)]		All nitrogen: catalytic decomposition, combined removal of NH_3/H_2S
NH_3		removed by aqueous scrubber[2)] removed to specification[2,3)]	Selective oxidation Under development
HCN		active coal bed[2)] possibly preceded by hydrolysis to NH_3[3)] Specifications are met.	
Sulphur compounds	total sulphur <1 ppmV[2)] < 10 ppb total S[3)]		All sulphur: In-bed Calcium sorbents Metal oxide sorbents < 20 ppm
H_2S		ZnO guard bed[2)] in case of high sulphur loads a special removal step, *e.g.* Claus unit.	
COS		active coal bed[2)] possibly preceded by hydrolysis to H_2S[3)] Specifications are met.	

1) Gas turbine specifications are met when FT specifications are.
2) Boerrigter et al. (2002).
3) Assumed by Tijmensen (2002).
4) Bechtel (Bechtel 1996).
5) Hot gas cleaning was practiced in the Värnamo Demonstration plant, Sweden (Kwant 2001). All data on dry gas cleaning here is based on the extensive research into high temperature gas cleaning by Mitchell (Mitchell 1997; Mitchell 1998).
6) Bergman et al (Bergman et al. 2002).

In dry gas cleaning, residual contaminations are removed by chemical absorbents at elevated temperature. In the FT situation, hot gas cleaning has little *energy* advantages as the FT reactor operates at 200 – 250 °C and especially when additional compression is required (efficient compression requires a cold inlet gas). However, dry gas cleaning may have lower operational costs than wet gas cleaning (Mitchell 1998). In the present study three gas-cleaning trains are applied, as depicted in Figure 2-2.

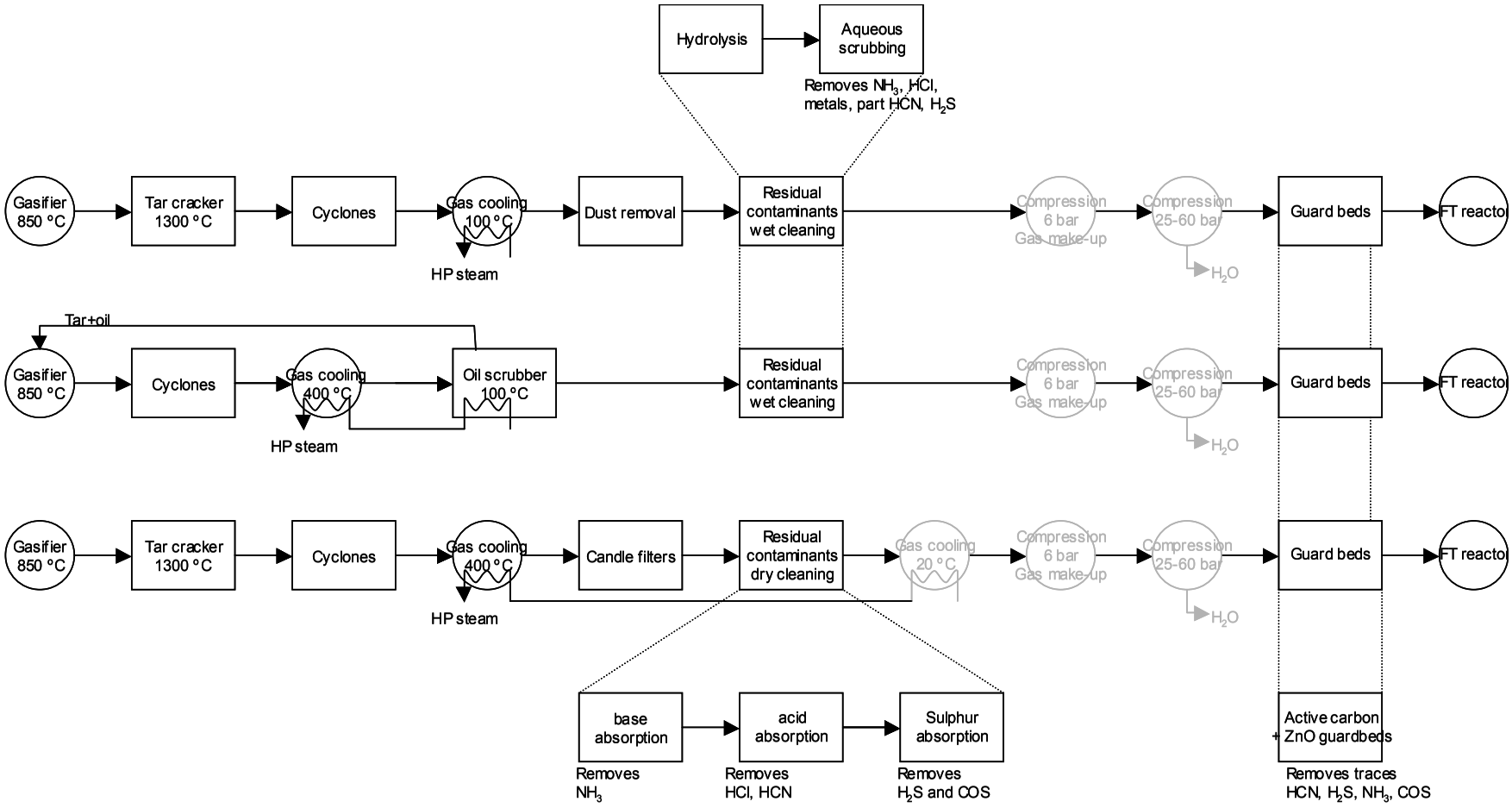

Figure 2-2. Three gas cleaning trains applied in this study. Top: tar cracking and conventional wet gas cleaning; middle: tar scrubbing and conventional wet gas cleaning; and bottom: Tar cracking and dry gas cleaning.

2.2.3 Gas conditioning

Early compression reduces the size of gas cleaning equipment, however, sulphur and chloride compounds condense when compressed and may corrode the compressor. Therefore, intermediate compression to 6 bar takes place after bulk removal and 60 bar compression after the guardbed.

The syngas can contain a considerable amount of methane and other light hydrocarbons, representing a significant part of the heating value of the gas. Reforming is optional, to convert these compounds to CO and H_2, driven by steam addition over a nickel catalyst. In the present study, autothermal reforming is applied, which is of simpler design than steam reforming, possibly cheaper (Katofsky 1993), and able to prevent carbon deposition (Tijmensen et al. 2002). Different reforming technologies yield different H_2/CO ratios, but this influence is small when processing bio-syngas.

The syngas produced by CFB gasification (especially at atmospheric pressures) generally has a low H_2/CO ratio. Although Fischer Tropsch selectivity is favourable at a low ratio, for high conversion effectively 2 H_2 are needed per CO. This can be realised by water gas shift with subsequent H_2 separation.

Water in the FT reactor does not poison the catalyst, but acts as inert, thereby decreasing the partial pressure of carbon monoxide and hydrogen. Therefore, prior to the FT reactor the syngas is cooled and water is knocked out.

The syngas from the gasifier contains a considerable amount of CO_2. After reforming or shifting, this amount further increases. CO_2 is no reactant in the FT reaction (assuming no shift reaction) but its presence negatively influences the C_{5+} selectivity. A physical absorption technology like PSA or ceramic membranes may be applied to remove CO_2. In PSA, both water and carbon dioxide are removed simultaneously.

2.3 FT synthesis and upgrading, and recycling

2.3.1 FT synthesis

In the FT reactor a high selectivity for liquids should be combined with a high conversion. The growth probability or selectivity, is largely catalyst dependent (Shell 2001) but depends furthermore on temperature, local partial pressures of reactants and inerts, and the applied FT technology (Dry 1981; Schulz 1999; Tijmensen et al. 2002). For example, α increases with decreasing H_2/CO ratio, decreasing reaction temperature and increasing pressure, and α is higher for Ru and Co catalysts relative to Fe catalysts (Bartholomew 1990). Conversion depends especially on availability of reactants and the reactor size.

There are three main kinds of FT reactors: the fixed bed reactor, the fluidised bed reactor and the slurry phase reactor. Of all these reactor types commercial versions exist (Shell, Sasol). Comparisons between the reactor types are broadly found (Bartholomew 1990; Sie and Krishna 1999; Tijmensen et al. 2002) and summarised in Table 2-2. Some parties favour the fluidised bed, and some the slurry phase reactor.

Table 2-2. Factors determining reactor choice.

	Multi-tubular fixed bed	Fluidised bed	Slurry Bubble column
Construction	Complex[1)]	-	Simple[1)]
Heat exchange	limited conversion per pass and large particle size necessary to obtain even temperature profile[2)]	Excellent heat transfer results in isothermal conditions[2)]	Excellent heat transfer results in isothermal conditions[3)]
Solids separation	not needed[2)]	Filtering may be necessary[2)]	Filtering is necessary[2)]
Gas-liquid separation	-	Foam formation occurs[2)]	-
Reactant distribution	Difficult[2)]	Automatically lateral mixed[2)]	Automatically lateral mixed[2)]
Catalyst settling or agglomeration	-	-	Too low velocities may lead to concentration gradient, insoluble materials may deposit on catalysts and hamper proper suspension[2)]
Catalyst deactivation	The top region (where the syngas enters) acts as a sulphur trap, whereas in the bottom region almost no sulphur is found but water inhibition plays a more important role.	-	Sulphur poisoning and water inhibition have evenly effect over the reactor (continuous circulation), leading to 1.5 – 2 times higher conversion loss than in fixed bed. High costs of cobalt catalysts necessitates a more effective sulphur removal[4)]
Loading/unloading catalyst	difficult, shutdown necessary[2)]	on-line[2)]	on-line[1,2)]
Scale-up	relatively straight forward by multiplying tubes[2)]	More complex and more costly, empirical scale-up through demonstration stage[2)]	Complex, but rational scale-up possible[2)]
Maximum capacity	-	-	2.5 – 6 times larger than fixed bed (if reactor size is weight limited)[1,2)] 2500 tonne/day (about 20,000 bbl/day) may be achieved[2)]

1) Jager (1997).
2) Sie and Krishna (1999).
3) Van der Laan (2000).
4) Espinoza (1999).

2.3.2 FT offgas

Conversion in the Fischer-Tropsch reactor is limited to a certain extent, depending on catalyst type and reactor size and technology. The reactor product stream thus contains unreacted carbon monoxide and hydrogen in addition to the FT products. The C_{5+} products are easily separated by a condensation step and sent to the recovery and upgrading section. It is in general not economic to recover the very small fraction of C_4 (Bechtel 1998).

To maximise the production of FT liquids, the offgas containing unreacted H_2 and CO and produced lower alkanes can be recycled to the entrance of the reactor. To limit the inert build up, the recycled volume is maximally equal to the main stream. The recycle can contain a reformer to reconvert C_{1-4} back into syngas, and a shift reactor. Furthermore the FT off gas may be recycled to the catalytic or thermal (1200-1400 °C) tar cracker, which will work as reformer. This latter option is only sensible when the gasification pressure is the same as the FT pressure. Recycling the off gas to the gasifier may be not sensible, because the gasification temperature is too low to reform methane. With higher once through conversions in future, the recycling option could become obsolete.

Instead of maximised fuel production, the system can also be optimised towards combined fuel and electricity production. In this case the syngas passes ‘once through’ the FT reactor. The FT offgas is not recycled but completely purged to a combined cycle for electricity production. Investments are

lower and presence of inert is less problematic. Co-production of FT liquids and electricity may combine a high overall efficiency with lower investment costs.

The hydrocarbon recovery plant recovers the C_{5+} naphtha, distillate and wax fractions for further processing. First the C_{5-10} fraction is separated from the heavier products. Although the C_{5-10} fraction cannot be applied as green petrol (too low octane number), it may have a high market value in the future as it is suitable for FCV vehicles with onboard reformer/POX, and can serve as raw material in green plastics production.

The waxy part of the raw synthesis product is selectively hydrocracked to the desired middle distillate products. Simultaneously, the product is isomerised to improve the cold flow properties, and subsequently fractionated in a conventional distillation column (Shell 2001).

3 System selection

In Chapter 2 the building blocks for Fischer-Tropsch Diesel production have been discussed. Given all the options for gasification, gas cleaning and conditioning, fuel synthesis and upgrading, and power generation, many routes to produce FT Diesel can be imagined. Figure 3-1 depicts the many possible combinations and choices.

Not all combinations may be sensible from efficiency or economic viewpoint. Besides, not all sensible combinations can be modelled due to time constraints. Following considerations lead to a limited set of configurations:

- The research focus lays on maximised FT production, especially by applying optimised once through conversion. In some concepts a recycle is applied.
- The gasifier chosen in this study is a direct fired circulating fluidised bed (TPS like). Gasification can be at atmospheric, intermediate, or high pressure (*i.e.* 1,6, and 25 bar, respectively). The gasification pressure is limited to 25 bar because of constructional and engineering considerations. The gasification medium is air, oxygen enriched air, or pure oxygen. The atmospheric air blown gasifier serves as starting point.
- Gas cleaning consists basically of tar removal and residual contaminations removal. Three configurations are considered: Tar removal by thermal cracking, followed by wet or dry residual cleaning, or tar removal by oil scrubbing preceding wet residual cleaning only. Wet cleaning to remove residual contaminations consists of several water-based scrubbers. In dry cleaning, residual contaminations are removed by chemical absorbents at elevated temperature. Although dry high temperature gas cleaning is not expected to have energetic advantages, as compression necessitates cooling anyway, and as the FT reaction itself runs at an intermediate temperature. However, there may be operational advantages.
- At atmospheric pressures an oil scrubber would remove BTX to a small extent only. Therefore, after atmospheric gasification only a tar cracker is applied for tar removal. After elevated pressure gasification, both tar cracking and oil scrubbing are evaluated. Heat recovery before tar removal – in the case of oil scrubbing – is possible as long as the temperature remains above the tar condensation or dew point (~400°C).
- After atmospheric gasification the feed gas is compressed to intermediate pressures between tar cracker and residual contaminations removal, and further to FT pressure after gas cleaning. Elevated pressure gasification concepts need only one compression step: after the gas cleaning train.
- Water has to be removed, because as inert it hinders the FT synthesis. Removal is usually done by condensation through compression and cooling. The sensitivity for inert presence is analysed by incorporating a model in which water is not removed. Processes for CO_2 removal, such as Selexol or amine type processes, also remove water.
- Reforming is incorporated in the recycle stream of recycle options and may be applied in once through options. The full recycle with tar cracker lacks a reformer, as all alkanes present will be reformed in the tar cracker. The reformer is operated at 6 or 25 bar, higher pressures negatively affect the equilibrium.
- Hydrogen needs to be separated for hydro cracking in the upgrading section. Water gas shift and separation can consecutively be done by a partial shift reactor followed by a pressure swing adsorption unit, or concurrently by a ceramic membrane. Hydrogen separation takes place directly preceding the FT reactor, in a side stream to limit the dimensions of reactors.
- Dry gas cleaning is always combined with ceramic membranes to represent advanced technological options; wet gas cleaning is followed by shift+PSA and represents conventional technology.
- It is assumed that CO_2 concentrations in the feed gas exceeding 10 % justify CO_2 removal, as investment costs for chemical absorption increase inversely with decreasing CO_2 concentration (Hendriks 1994). The combined physical/chemical absorption unit is placed just before the FT reactor.
- Both once through and recycle options are evaluated to investigate both optimisation towards FT products, and combined FT and electricity production. The short recycle option basically produces hydrogen from the off gas instead of the feed gas, to keep the hydrogen/carbon monoxide ratio below the consumption ratio of these compounds over reactor length and for hydro cracking purposes. A few full recycle concepts are included in which the tar cracker acts as a reformer for lower alkanes. It must be remarked that in these cases a huge clean gas stream is pushed through the gas cleaning section. To avoid accumulation of inert, recycle options are only evaluated for

syngas from oxygen blown gasifiers. The full recycle is operated with both gasifier and FT reactor at 25 bar. The recycle is maximally equal to the feed stream.

- Both fixed bed and slurry Fischer Tropsch reactor are analysed. Slurry reactors are CSTR type rather than plug flow type, selection towards C_{5+} products is a function of the H_2/CO ratio at the reactor outlet, as described in § B.2, and staged hydrogen injection is not necessary.
- As the gas cleaning will be good enough to preserve the catalyst, cobalt catalysts are preferred over iron catalysts, as cobalt catalyst gives a higher conversion.
- The FT reactor is operated at 60 bar when the feed gas stream contains much inert. The sensitivity towards lower FT pressures (40 and 25 bar) in N_2 lean once-through options will also be examined. In the short and full recycle options, the FT reactor is operated at 25 bar.
- The FT reactor produces a liquid flow with C_{5+} products and a gaseous flow with unreacted CO, H_2, and C_{1-4} products. All C_5 and heavier go to the recovery and upgrading section. The FT off gas may be partly recycled and the rest is directed to electricity production.
- Electricity is produced from off gas by a combined cycle, or a boiler followed by a steam turbine. The choice for either configuration depends on the quality of the off gas (Hamelinck et al. 2001).
- For scale advantages it could be attractive to co-feed natural gas to the FT plant, or co-fire natural gas in the gas turbine. The FT Diesel would become less 'green' but the economies of the green part would become better, since the plant can be much larger. However, this project focuses on the abilities for "stand alone" FT production.
- The heat produced on various places in the plant is carefully matched with the heat demand at other places in the plant. Leftover heat is used to raise steam of different qualities, which is eventually used for power production through a steam turbine. There are no net heat streams to or from the plant.

These considerations lead to a selection of promising biomass to FT conversion concepts, summarised in Table 3-2 and envisioned in Figure 3-1 through Figure 3-4. The main variations between these concepts stem from the different gasifier configurations (Table 3-1) and the gas cleaning train (Figure 2-2). Some concepts represent the long term, while others are feasible at present with state of the art technology.

Table 3-1. Gasifier configurations and corresponding FT reactor pressure, applied in this study.

	Gasification medium	Gasification pressure		
		1 bar	6 bar	25 bar
Once through				
	air	60	-	-
	enriched[1)]	40[2)]/60	40[2)]/60	40[2)]/60
	oxygen	25[2)]/40[2)]/60	25[2)]/40[2)]/60	25[2)]/40[2)]/60
Short recycle				
	oxygen	25	25	25
Long recycle				
	oxygen	-	-	25

1) Consists of 80 % oxygen and 20 % air.
2) For sensitivity analysis.

The technical and economic performance of these concepts are analysed in the following section, for small (80 MW_{th} input), intermediate (400 MW_{th}), and large (2000 MW_{th}) scale. The FT reactor (catalyst) will be modelled performing state of the art, with a sensitivity analysis towards improved catalysts.

Table 3-2. Selected Fischer-Tropsch production concepts.

	Gasification		Tar removal Gas cleaning		Reforming Shift and H_2 separation	FT
Once through concepts						
I	1 bar	air	tar cracker 1300 °C	HX 100 °C wet cleaning	(ATR reforming) side stream: shift reactor + PSA	60 bar
II		enriched				
III		oxygen				
IV	6 bar	enriched				
V		oxygen				
VI	25 bar	enriched				
VII		oxygen				
VIII	1 bar	air		HX 400 °C dry cleaning	(ATR reforming) side stream: Ceramic Membrane	
IX		enriched				
X		oxygen				
XI	6 bar	enriched				
XII		oxygen				
XIII	25 bar	enriched				
XIV		oxygen				
XV	6 bar	enriched	HX 400 °C Oil scrubber 100 °C	wet cleaning	(ATR reforming) side stream: shift reactor + PSA	
XVI		oxygen				
XVII	25 bar	enriched				
XVIII		oxygen				
Short recycle concepts (to reformer)						
XIX	1 bar	oxygen	tar cracker 1300 °C	HX 100 °C wet cleaning	In recycle: ATR reformer shift reactor + PSA	25 bar
XX	6 bar					
XXI	25 bar					
XXII	1 bar			HX 400 °C dry cleaning	In recycle: ATR reformer ceramic membrane	
XXIII	6 bar					
XXIV	25 bar					
XXV	6 bar		HX 400 °C Oil scrubber 100 °C	wet cleaning	In recycle: ATR reformer shift reactor + PSA	
XXVI	25 bar					
Long recycle concepts (to tar cracker)						
XXVII	25 bar	oxygen	tar cracker 1300 °C	HX 100 °C wet cleaning	No reformer side stream: shift reactor + PSA	25 bar
XXVIII				HX 400 °C dry cleaning	No reformer side stream: ceramic membrane	
XXIX			HX 400 °C Oil scrubber 100 °C	wet cleaning	No reformer side stream: shift reactor + PSA	

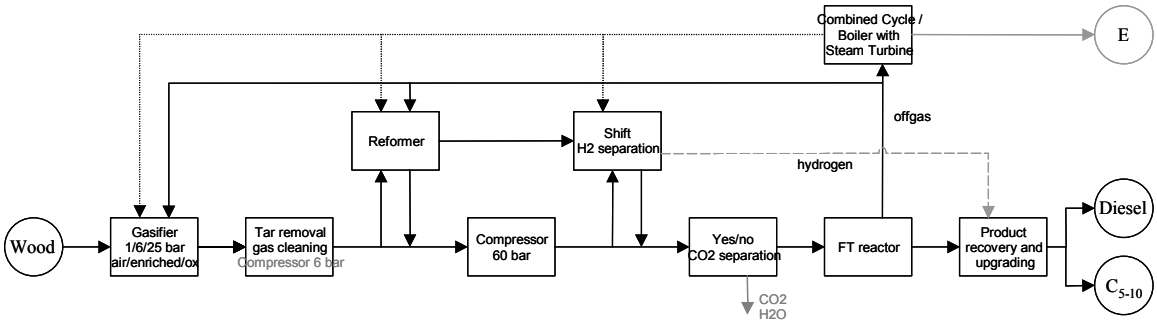

Figure 3-1. Summary of possible process configurations for FT production.

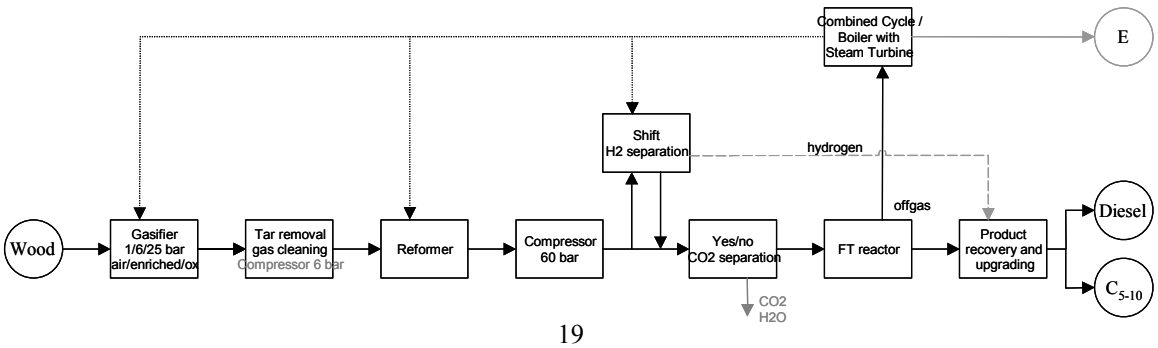

Figure 3-2. Once through concepts.

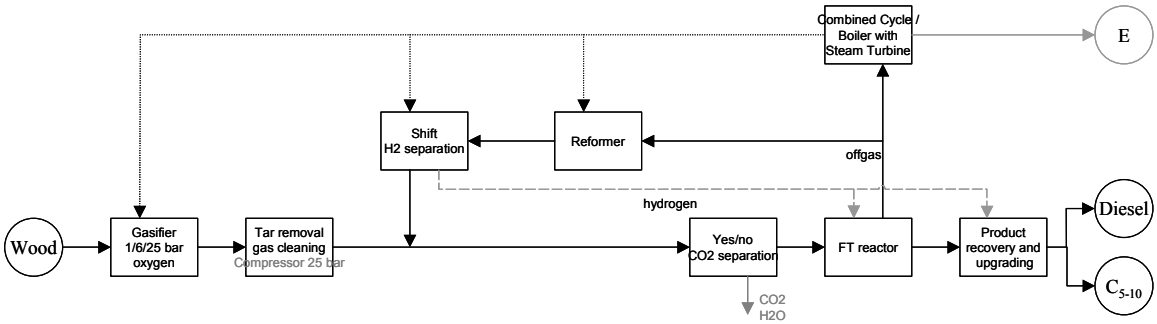

Figure 3-3. Short recycle concepts.

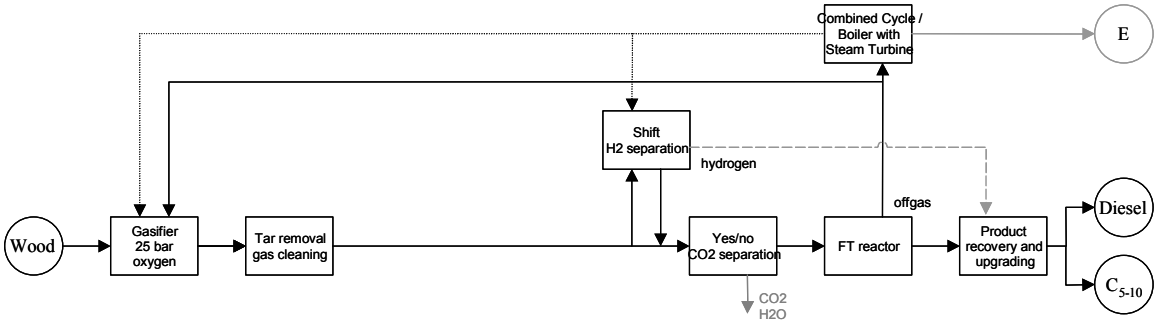

Figure 3-4. Full recycle concepts.

4 System modelling

To evaluate the chosen configurations, one general flowsheet is built in Aspen Plus, following the scheme in Figure 3-1. By "opening or closing valves", the different system configurations arise, allowing efficient and quick modelling, and direct comparisons. Modelling assumptions are summarised in Table D-1 in Annex D. The gasifier, reformer and gas turbine deliver heat, whereas the dryer, gasifier, reformer, and shift require steam. The supply and demand of heat is added to or drawn from the steam turbine, such that the surplus heat is turned into electricity.

Using a "gasifier databank", different gasifiers can be 'plugged' into the plant (see Table 4-1). The pressure and oxygen purity determine the amount of oxidation medium and steam. Relations from literature determine the yields of methane and heavier components, the other products follow from the water gas shift equilibrium, see footnotes under Table 4-1. It is also possible to choose between differently performing FT reactors (catalysts) and different combined cycles, in aid of the sensitivity analysis.

Table 4-1. Mass flows for circulating fluidised bed gasifier (850 °C), for different media and pressures. Input scale is 440 MW HHV biomass (421 MW LHV). It is assumed that the results may be scaled linear towards other capacities.

Oxidative medium[1)]		**Air**			**Enriched air**			**Oxygen**	
pressure (bar)	**1.3**	**6**	**25**	**1.3**	**6**	**25**	**1.3**	**6**	**25**
In (kg/s)									
Biomass[2)]	26.5	26.5	26.5	26.5	26.5	26.5	26.5	26.5	26.5
Oxidative medium[3)]	32.5	30.2	28.0	8.50	7.76	7.09	6.79	6.21	5.67
Steam[4, 5)]	0	0	0	14.7	13.1	11.6	11.0	9.68	8.49
Out[6)] (kg/s)									
H_2	0.661	0.559	0.462	0.808	0.663	0.533	0.791	0.648	0.520
CO	10.7	9.67	8.62	5.84	5.16	4.51	6.74	5.98	5.25
CO_2	13.3	13.6	14.0	19.5	19.1	18.9	18.3	18.0	17.9
H_2O	5.32	5.12	4.88	17.5	15.9	14.5	13.9	12.7	11.5
CH_4	1.76	2.28	2.78	2.32	2.90	3.42	2.34	2.91	3.44
C_2H_4	0.328	0.310	0.29	0.315	0.293	0.272	0.279	0.260	0.242
C_6H_6	0.365	0.346	0.33	0.351	0.326	0.303	0.311	0.289	0.270
N_2	25.3	23.5	21.8	1.77	1.64	1.51	0.271	0.268	0.266
Tar[7)]	0.338	0.338	0.338	0.338	0.338	0.338	0.338	0.338	0.338
Ash	0.428	0.428	0.428	0.428	0.428	0.428	0.428	0.428	0.428
C	0.551	0.551	0.551	0.551	0.551	0.551	0.551	0.551	0.551
Total (kg/s)	59.0	56.7	54.5	49.7	47.4	45.2	44.3	42.4	40.6

1) Air at 450 °C, Enriched air (80 % O_2) at 200 °C, oxygen (99.5 % O_2) at 200 °C.

2) Feedstock is willow wood: moisture 15 wt %, ash 1.9 wt % d.b., C 49.9 wt % d.a.f., H 6.11 wt % d.a.f., O 42.9 wt % d.a.f. N 0.62 wt % d.a.f. HHV = 19.88 MJ/kg d.a.f. (Phyllis).

3) It is assumed that carbon conversion is 95 % and heat loss is 2 % of the thermal input. The amount of medium is adjusted such that the desired temperature is reached.

4) Steam at 450 °C.

5) Steam replaces the heat capacity of nitrogen.

6) The water-gas shift reaction is assumed to be at equilibrium. Tar yield from gasification is assumed to be 1.5 wt % of the dry feed. Ethane and benzene concentration is the product gas are assumed to be 0.5 and 0.2 vol % respectively. The methane yield is assumed to be in pseudo-equilibrium (Liinanki et al. 1985):

$$K \cdot \log\left(\frac{p_{H_2}^{1.3}}{p_{CH4}}\right) = 3.0 \times 10^4 \exp\left(\frac{-1.17 \times 10^4}{T}\right)$$

7) Tar is defined as $C_{14}H_{10}$, represented in Aspen Plus by phenanthrene.

4.1 FT modelling

Selection and conversion in FT synthesis are a function of feed stream composition, pressure, temperature, catalyst and reactor type and size. Figure 4-1 depicts how the influence of various parameters is incorporated in the Aspen Plus model.

For high C_{5+} selectivity, the hydrogen/carbon monoxide ratio should be kept below the consumption ratio of these compounds. For the solid bed reactor, which is of plug flow type, this inevitably leads to a decrease of the actual ratio during conversion, such that selectivity and conversion rate change over the reactor length. In the slurry reactor (CSTR type) the bulk is assumed totally mixed, so that every local ratio equals reactor exit condition. In reality the gas phase behaves in plug flow.

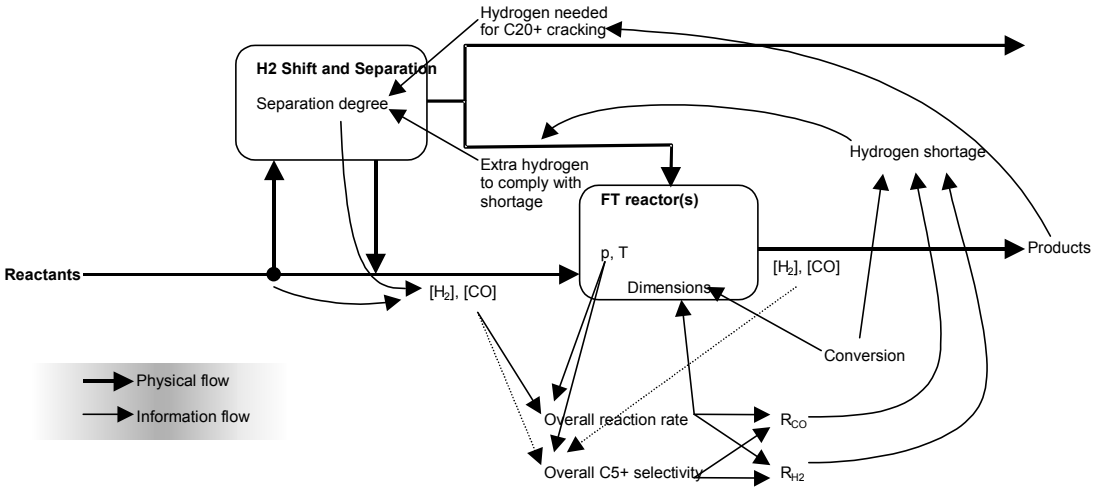

Figure 4-1. Schema for incorporation of selectivity and conversion in Aspen Plus model. The dashed line illustrates that the C_{5+} selectivity is influenced by either entrance or exit concentrations, depending on the reactor type: solid bed respectively slurry phase. It is also possible to fix the reactor dimension and calculate the conversion.

4.1.1 Selectivity

In literature, relevant relations for the selectivity depending on these parameters are mostly described in a qualitative way. Selectivity e.g. decreases with hydrogen content, decreases with temperature, and increases with pressure. Assuming that linear relations hold (within boundaries) for selectivity dependency on temperature, ratio, and syngas pressure, the following general relation holds:

$$S_{C5+} = C_1 + C_2 \cdot T + C_3 \cdot \frac{[H_2]}{[CO]} + C_4 \cdot ([H_2] + [CO]) + C_5 \cdot p_{Total}$$ Equation 4-1

A least sum of squares fit with literature data (Dry 1981; den Uil et al. 2001) gives Equation 5-2.

$$S_{C5+} = 1.696 - 0.00241 \cdot T - 0.0876 \cdot \frac{[H_2]}{[CO]} + 0.181 \cdot ([H_2] + [CO]) + 0.00781 \cdot p_{Total}$$ Equation 4-2

In the computer model, the selectivity for slurry reactors is calculated from exit conditions, and for solid bed from entrance conditions, change of selectivity along the solid bed is assumed negligible. The product distribution in the relevant range of C_5-C_{20} is assumed to obey the ASF distribution. The relation between α And S_{C5+} can then be approached (see Annex C.2) by:

$$\alpha = 0.75 - 0.373 \cdot \sqrt{-\log(S_{C5+})} + 0.25 \cdot S_{C5+}$$ Equation 4-3

Now the computer model can calculate the $C_{1\text{-}19}$ from the ASF formula. The fraction $C_{4\text{-}}$ is redistributed to 74 mol% C_1, 16 mol% C_2, 6 mol% C_3, and 4 mol% C_4 (van der Laan and Beenackers 2000). C_{20+} is expressed in weight fractions C_{20} and C_{32}, by first calculating the average molecular mass (see Annex C.4). H_2O, CO_2, and N_2 entering the reactor are considered inert, thus decreasing the partial pressures of the reactants. Any hydrocarbons entering the reactor (after recycle) are inert as well and will not grow to longer chains.

4.1.2 Conversion

The reaction rate determines the conversion extent or the reactor size. Many different relations for reaction rate are found in literature, for different catalysts and sometimes incorporating water inhibition. Yates and Satterfield (1991) extensively reviewed Fischer-Tropsch over cobalt kinetic studies and found the following form to describe best the reaction:

$$-R_{CO} = \frac{a \cdot p_{CO} \cdot p_{H_2}}{(1 + b \cdot p_{CO})^2}$$ Equation 4-4

with $-R_{CO}$ = Carbon monoxide consumption rate in mol/(s·kg_{cat})

p_{CO}, p_{H2} = partial pressure of CO or H_2, in bar

The computer model calculates the reaction rate in the slurry reactor from the exit conditions. The rate in solid bed reactors is interpolated between rates at reactor entrance and exit. Kinetic parameters used for modelling in the present study are given in Table 4-2.

Table 4-2. Kinetic modelling parameters[1)].

a	(mol/(s·kg_{cat}·bar^2))	$k_0 \cdot \exp\left(\frac{-E_{A,reaction}}{R \cdot T}\right)$	
b	(in 1/bar)	$k_1 \cdot \exp\left(\frac{-\Delta H_{ads}}{R \cdot T}\right)$	
R	(J/(K·mol))	8.31	
T	(K)		
		Solid bed reactor	**Slurry reactor**
$-E_A$	(kJ/mol)	-68 [2)]	-115 [2)]
k_0	(mol/(s·kg_{cat}·bar^2))	$1.2 \cdot 10^5$	$1.0 \cdot 10^{10}$
$-\Delta H_{ads}$	(kJ/mol)	192 [3)]	192 [3)]
k_1	(1/bar)	$3.5 \cdot 10^{-23}$	$3.5 \cdot 10^{-23}$
d	($kg_{cat}/m^3_{reactor}$)	1200 [4)]	600 [4)]

1) Yates and Satterfield give values for the kinetic rate constant and adsorption rate constant in Equation 5-4, for Co catalysts in slurry at 220 and 240°C (Yates and Satterfield 1991, Table V). These values may not be used to calculate the E_A and ΔH_{ads}, because of the small temperature range. However, with known E_A and ΔH_{ads} (see notes 2 and 3), k_0 and k_1 can be calculated.

2) The apparent activation energy for the FT reaction depends strongly on diffusion limitation, which in turn depends on catalyst formulation and structure. When the catalyst particles are very small (slurry reactor), the reaction is not diffusion limited and the measured activation energy is about 100 kJ/mol for Co catalysts (Yates and Satterfield 1991). Under diffusion limitation (large pellets in solid reactor) the apparent activation energy is about half of the E_A without diffusion limitation (small pellets in slurry reactor). Values from Iglesia et al. (Iglesia et al. 1993, Table VII).

3) (van der Laan 1999, Figure 2.1).

4) Pure catalyst weights 2 kg/l, solid bed has a porosity of ~40 vol %. Concentration in the slurry phase is assumed 30 vol % or 0.6 kg/l (Sie and Krishna 1999).

5 Economic calculations

Concurrent with the Aspen Plus technical performance calculations, an economic evaluation is carried out for the each concept. The Fischer-Tropsch Diesel production costs are calculated by dividing the total annual costs of a system by the produced amount of fuel. The total annual costs consist of annual capital investments, operating and maintenance (including maintenance, consumables, labour, waste handling), biomass feedstock and electricity supply / demand (fixed power price). Relevant parameters to calculate these costs and the resulting FT fuel costs, are given in Table 5-1.

Table 5-1. Parameters for the economic evaluation[1)].

Interest rate	10 %
Economical lifetime	15 years
Technical lifetime	25 years
Operational costs[2)]	
Maintenance	3 % of Total Capital Investment
Labour	0.5 % at 400 MW HHV input decreasing with scale (R = 0.25)
Dolomite	8.7 M€ at 400 MW HHV input linear with scale
Gas cleaning	wet: 0.5 % of TCI, dry: 0 %
Insurance	0.1 % of TCI
Biomass[3)]	3 $€_{2002}/GJ_{HHV}$ (short term) - 2 $€_{2002}/GJ_{HHV}$ (long term)
Electricity	0.03 €/kWh
Annual load	8000 hours

1) Most of these parameters come from Hamelinck et al. (2001).

2) A breakdown of operational costs for BIG/CC was applied by Faaij et al. (1998): Maintenance costs are 3 % of Total Capital Investment. Operation costs depend on labour (Dierendonck 1990), catalyst and chemicals consumption, residual streams disposal, and insurance. Labour costs decrease with scale and are 0.5 % of the TCI for a 400 MW HHV input. Dolomite costs are 50 €/tonne, dolomite use is 0.3 kg/kg clean dry wood. At 400 MW HHV input the annual biomass throughput is 0.56 Mtonne. Wet cleaning costs are in waste water treatment and NaOH consumption, while dry residuals are assumed suitable for construction. Insurance is 1 % of annual depreciation.

3) Prices of delivered cultivated energy crops and forrest thinnings in Western Europe amount currently 3-5 $€/GJ_{HHV}$, United States 2.3-3.3 $€/GJ_{HHV}$, and at some Latin American locations even lower costs are possible 1.2 - 2 $€/GJ_{HHV}$ (Bauen 2001; Perlack et al. 1995). Improved crops and production systems are expected to bring the biomass price to the 1.6-2.1 $€/GJ_{HHV}$ range (United States, Turkenburg 2000).

The total capital investment, or TCI, is calculated by *factored estimation* (Peters and Timmerhaus 1980; Hamelinck et al. 2001, and Annex F to this report), based on known costs for major equipment as found in literature or given by experts. The uncertainty range of such estimates is up to ± 30 %. Usually the TCI follows from multiplying the total purchased equipment costs by a factor to yield the Inside Battery Limit (ISBL)[2], adding the costs of the Outside Battery Limit (OSBL) to yield the direct costs, adding indirect costs to yield the fixed capital investment, and finally adding working capital and start-up costs. However, the advised method cannot be followed entirely as base equipment costs in literature may be anywhere between f.o.b.[3] and total installed capital, and often a specific overall installation factor is given to yield the TCI of that piece of equipment.

Therefore the TCI is first calculated for each separate unit, and later all unit TCIs are added up. The unit TCI depends on the size of that unit (which follows from the Aspen Plus modelling), by scaling from a known scale (see Table 5-2). Various system components have a maximum size, above which multiple units will be placed in parallel. Hence the influence of economies of scale on the total system costs decreases. This aspect is dealt with by assuming that the base investment costs of multiple units are proportional to the cost of the maximum size: the base investment cost per size becomes constant. After scaling, the TCI of each component is found by multiplying the scaled base cost by an installation factor.

The given numbers must be used carefully, as the base equipment costs in literature are often ill defined. Also the applied percentages for additional costs differ greatly between studies. Of course these percentages depend on the specific location: OSBL costs in industrial areas may be lower since

[2] *Inside Battery Limits* only deals with the purchase and installation of process equipment, piping, instrumentation, controls, process buildings, etc. *Outside Battery Limits* includes utilities such as power distribution, steam plants, instrument air systems, sewers, waste water treatment, tankage, cooling towers, control buildings, land, etc.

[3] *Free on board*: the price of a traded good after loading onto a ship but before shipping, thus *not* including transportation, insurance, and other costs needed to get a good from one country to another (Bannock et al. 1998); cost of equipment ready for shipment from supplier.

the necessary infrastructure is readily available. E.g. a Middle East location will differ greatly from a Western Europe location and OSBL are higher. All the factors to calculate TCI from purchased equipment are expected to be very dependent on scale, as ISBL relatively decreases with scale because of increasing process integration and OSBL decreases because certain off-sites have a fixed same scale for both small and large plants. However, no literature was found on this subject, therefore the influence of scale and/or location on the overall installation factor is generally not incorporated in the present study. Only where the composition of the installation factor (in direct, indirect, working capital and start-up) is known, the direct costs part of the overall installation factor is decreased with scale (average R-factor is 0.82, derived from Faaij et al. 1998).

Table 5-2. Costs of system components in M€$_{2002}$ [1].

Unit	Base cost	Scale Factor	Base Scale	Overall installation factor [25]	Maximum Size [26]
Pre-treatment [2]					
Conveyers [3]	0.41	0.8	33.5 wet tonne/hr	2.0	110
grinding [3]	0.48	0.6	33.5 wet tonne/hr	2.0	110
storage [3]	1.16	0.65	33.5 wet tonne/hr	2.0	110
dryer [3]	8.5	0.8	33.5 wet tonne/hr	2.0	110
iron removal [3]	0.43	0.7	33.5 wet tonne/hr	2.0	110
feeding system [3,4]	0.48	1	33.5 wet tonne/hr	2.0	110
Gasification System					
CFB gasifier [5]	44.3	0.7	68.8 dry tonne/hr	1.69	75
Air separation unit 99.5 % O_2 [6]	27.9	0.75	576 tonne/day O_2	1.3	3200
Oxygen compressor [7]	18.1	0.85	13.2 MW_e	1.86	-
Gas Cleaning					
Tar cracker [3]	3.6	0.7	34.2 m^3/s gas	2.0	52
Oil scrubber [8]	1.64	0.7	14.7 m^3_{NTP}/s gas	1	-
Cyclones [3]	3.0	0.7	34.2 m^3/s gas	2.0	180
High-temperature heat exchanger [9]	8.1	0.6	138.1 MW_{th}	1.84	-
Particle filters [3]	1.9	0.65	12.1 m^3/s gas	2.0	64
Scrubbers [3]	3.0	0.7	12.1 m^3/s gas	2.0	64
Dry gas cleaning [10]	35.8	1.0	74.1 m^3/s gas	1.86	-
Guard beds (ZnO + active C) [11]	0.024	1.0	8.0 m^3_{NTP}/s gas	3	
Syngas Processing					
Compressor [12]	12.9	0.85	13.2 MW_e	1.86	-
Autothermal reformer [13]	31.1	0.6	100 m^3_{NTP}/s	2.3	-
Shift reactor [14]	12.2	0.65	8819 kmol CO+H_2/hr	1.81	-
PSA units A+B [15]	32.6	0.7	9600 kmol feed/hr	1.69	-
Ceramic membrane [16]	25.2	0.8	17 tonne H_2/hr	1	-
Selexol CO_2 removal [17]	63.0	0.7	9909 kmol CO_2/hr	1	-
Fischer-Tropsch Production					
Solid bed gas phase FT 60 bar [18]	25.3	1	208 m^3	1.30	-
Slurry phase FT 60 bar [19]	36.5	0.72	362 m^3	1	365
Product upgrading [20]	233	0.7	286 m^3_{FT}/hr	1	
Power Isle [21]					
Gas turbine + HRSG [3,22]	22.0	0.7	26.3 MW_e	2.0	-
Steam turbine + steam system [3,23]	5.9	0.7	10.3 MW_e	2.0	-
Expansion turbine [24]	5.0	0.7	10.3 MW_e	2.0	-

1) Annual GDP deflation up to 1994 is determined from OECD (1996) numbers. Average annual GDP deflation after 1994 is assumed to be 2.5 % for the US, 3.0 % for the EU. Cost numbers of Dutch origin are assumed to be dependent on the EU market, therefore EU GDP deflators are used. 1 €$_{2002}$ = 0.88 US$$_{2002}$.

2) Total pre-treatment approximately sums up to a base cost of 11.46 €$_{2002}$ at a base scale of 33.5 tonne wet/hour with an R factor of 0.79.

3) Based on first generation BIG/CC installations. Faaij et al. (1995) evaluated a 29 MW_e BIG/CC installation (input 9.30 kg dry wood/s, produces 10.55 m_{NTP}^3 fuel gas/s) using vendor quotes. When a range is given, the higher values are used (Faaij et al. 1998). The scale factors stem from Faaij et al. (1998).

4) Two double screw feeders with rotary valves (Faaij et al. 1995).

5) Direct costs are 29.74 MUS$$_{1991}$ for a 1650 dry tonne/day input IGT gasifier, R = 0.7 (Williams et al. 1995). Maximum input is 400 MW_{th} HHV (Tijmensen et al. 2002).

6) Various cost numbers with a large distribution are found in literature (van Dijk et al. 1995; Williams et al. 1995; Larson et al. 1998; Bechtel 1996; Tijmensen et al. 2002), see Annex A. Here the cost number from Tijmensen (2002) is used because it bests fits the real oxygen cost price (Kirschner 1999): Direct costs are 27.9 M€$_{2002}$ for a base scale of 576 tpd oxygen of 99.5 % purity, 1 bar, the overall installation factor is constant 30 %, R = 0.75. It is assumed that oxygen of 95 % purity is 5 % cheaper; oxygen of 80 % purity for the scale considered here, will always be mixed from 95% O_2 and air, and thus barely has cost advantages. The highest maximum scale found to date is 3200 tpd (Shell Bintulu, see Wilhelm et al. 2001)

7) Because of explosion risk, oxygen is compressed in a non-lubricated compressor, which is 1.4 times more expensive than lubricated compressors (Livingston 1993).

8) OLGA technology (Bergman et al. 2002). Cost numbers are assumed by the authors.

9) High temperature heat exchangers following the gasifier and (in some concepts) at other locations are modelled as HRSG's, raising steam of 86 bar/510 °C. A 39.2 kg steam/s unit (i.e. transferring 138.1 MW_{th}) costs 6.33 $MUS\$_{1997}$ f.o.b., overall installation factor is 1.84 (Larson et al. 1998), which is assumed to consist of 22 % added direct costs (vary with scale) and 50 % added indirect costs (constant), see note 25.

10) Tijmensen (2002) assumes the f.o.b. price for Hot Gas Cleaning equipment to be 30 $MUS\$_{2000}$ for a 400 MW_{th} HHV input. This equals 74.1 m^3/s from a BCL gasifier (863 °C, 1.2 bar). There is no effect of scaling. The installation factor is taken from Faaij et al. (1998) but for 400 MW_{th} input scale (decreased added direct costs), see footnote 25.

11) For 427 MW_{th} LHV biomass input concepts, Tijmensen (2002) assumes that 1 wt% of the fuel S enters the ZnO bed, and two guard beds of 3 m^3 are required. These beds consist of 2300 kg steel, at a steel price 20 Dfl_{2000}/kg, so that each guard bed will cost 50.000 Dfl_{2000} f.o.b., or 150.000 Dfl_{2000} installed. Tijmensens 70 MW_{th} LHV (13 dry tonne/hour biomass, assumed gas yield 100 kmol/dry tonne) corresponds to a local gas flow of 8.0 m^3_{NTP}/s.

12) Katofsky (1993) assumes compressors to cost 700 $US\$_{1993}$ per required kW_{mech}, with an installation factor of 2.1. The relation used in the present study stems from the compressor manufacturer Sulzer quoted by Tijmensen (2002). At the indicated base-scale, total installed costs are about 15 % higher than assumed by Katofsky. Multiple compressors, for fuel gas, recycle streams, or hydrogen, are considered as separate units. The installation factor is taken from Faaij et al. (1998) but for 400 MW_{th} input scale (decreased added direct costs), see note 25.

13) Hamelinck (2001) based on Katofsky (1993) assumes 4.7 $MUS\$_{2001}$ with an overall installation factor of 2.3 for an 1390 kmol total/hour (8.8 m^3_{NTP}/s) unit, R=0.6, where Bechtel (1991-1994) gives a total capital investment of 20.52 $MUS\$_{1992}$ for an ATR unit of 364.7 MMscfd (120 m^3_{NTP}/s) gas feed, which is almost twice as costly. Here we assume the average between the two: 31.1 M€ total capital investment for a 100 m^3_{NTP}/s unit, R=0.6. The installation factor is assumed to consist of variable direct costs (53 % in base case) and fixed indirect costs (50 %), see note 25.

14) Cost numbers from Williams (1995) are used: 9.02 $MU\$_{1995}$ for an 8819 kmol $CO+H_2$/hr reactor, overall installation factor is 1.81 and R = 0.65. Hendriks (1994) writes that a reactor of 350.000 m^3_{NTP}/hr total gas ($CO+H_2$ is 93.3 %, this gives 14.6 Mmol/hr $CO+H_2$) costs 30 $MUS\$_{1994}$ installed (no R given) which corresponds with the line from Williams. Hendriks also gives an annual (sulphur tolerant) catalyst use of 0.7 $MUS\$_{1994}$, but considers this to be part of the operational costs. Numbers from Tijmensen (2002) are considered too low: 0.95 $MDfl_{2000}$ for direct costs plus engineering (add 30 % to obtain total capital investment) for a 2400 kmol/hr unit, R = 0.6.

15) PSA units (excluding the recycle compressor) cost 23 $MUS\$_{1993}$ for a 9600 kmol feed/hour throughput, R=0.7 (Katofsky 1993).

16) Membrane costs 68 $US\$_{1997}$/(kW/bar), but these costs are only 9 % of the total installed cost for a Hydrogen Separation Device. Investment costs stem from Parsons I&TG (1998). The economies of scale of the membrane surface are low because the required surface area is proportional to the throughput, this slightly influences the overall R factor of the complete HSD.

17) Costs for CO_2 removal through Selexol amounts 14.3 $MUS\$_{1993}$ fob (overall installation factor is 1.87) for an 810 kmol CO_2/hr unit, R=0.7 (Katofsky 1993) up to 44 $MUS\$_{1994}$ installed for a 9909 kmol CO_2/hour unit (Hendriks 1994). The value from Hendriks is assumed to be right, since his research into CO_2 removal is comprehensive.

18) FT Gas phase reactor operating at 40 bar costs 35 $MDfl_{2000}$ (direct costs + engineering) for a production of 100 MW_{FT} HHV (CO conversion = 150 mol/s, and $-R_{CO} = 6.03 \cdot 10^{-4}$ mol/s·kg_{cat}, thus reactor volume is 208 m^3 CNH), R = 1, there is no maximum scale, add 30 % to obtain total capital investment (Tijmensen et al. 2002). For operation at 60 bar and same volume, costs increase by 50 % (reactor with thin wall: thickness linear with pressure).

19) 19.995 $MUS\$_{1996}$ ISBL or 34.6 $MUS\$_{1996}$ "Total Costs" (includes offsites, HOService/Fee and contingency) here assumed to be total capital investment, for a FT slurry reactor train of *three* 362 m^3 reactors at 25.2 bar, each yielding 295 m^3_{FT}/day or 131 MW_{FT} (Bechtel 1996), the installation costs of one 362 m^3 reactor thus is 15.2 $M€_{2002}$. Operation at 60 bar makes the installation 2.4 times more expensive (see note 18).

20) Tijmensen (2002) assumes that a hydrocracking unit has installed costs (=Total capital investment, CNH) of 17.05 $MDfl_{2000}$ (8.22 $M€_{2002}$) per 2000 bbl/day or 11352 GJ_{C5+} per day (13.2 m^3/hr), but does not specify further, his source is confidential. According to Bechtel (1996) a wax hydrocracking plant of 475 Mlbs/hr of FT diesel+gasoline product (286 m_{FT}^3/hr) costs 71.5 $MUS\$_{1996}$ total capital investment, furthermore a distillate hydrotreater (29.1 $MUS\$_{1996}$), a naphtha hydrotreater (9.8), a naphtha reforming plant (49.3), a C_4 isomerisation plant (7.2), a C_5/C_6 isomerisation plant (11.2), a $C_3/C_4/C_5$ alkylation plant (41.2), and a saturated gas plant (9.2) are necessary, thus summing up to 233 $M€_{2002}$. Thus upgrading costs in Bechtel's study are 1.3 - 3.3 (at R = 1 - 0.7) times Tijmensen's values. Here we assume that the Bechtel numbers are right.

21) For indication: A complete Combined Cycle amounts about 830 $US\$_{1997}$ per installed kW_e. Quoted from (Solantausta et al. 1996) by (Oonk et al. 1997).

22) Scaled on Gas Turbine size.

23) Steam system consists of water and steam system, steam turbine, condenser and cooling. Scaled on Steam Turbine size.

24) Expansion turbine costs are assumed to be the same as steam turbine costs (without steam system).

25) Overall installation factor. Includes auxiliary equipment and installation labour, engineering and contingencies. Unless other values are given by literature, the overall installation factor is set 1.995 for a 70 MW_{th} scale (Faaij et al. 1998). This value is based on 33% added *direct costs* to installed equipment (instrumentation and control 5%, buildings 1.5% grid connections 5%, site preparation 0.5%, civil works 10%, electronics 7%, and piping 4%) and 50 % added *indirect costs* to direct costs (engineering 15%, building interest 10%, project contingency 10%, fees/overheads/profits 10%, start-up costs 5%). The added *direct costs* part of the overall installation factor decreases with scale (average R-factor is 0.82).

26) Maximum sizes from Tijmensen (2002), unless indicated otherwise.

6 System performance

6.1 Physical and economic performance

In Section 3 the most promising FT concepts were listed. Here we analyse the technical and economical performance of these concepts, in order to find the best configuration(s). Results for the first group of 7 once through concepts are given in Figure 6-1. These concepts apply a tar cracker and conventional wet gas cleaning, but no reforming. Obviously, the production of FT fuel becomes cheaper with increasing gasification pressure and oxygen purity. This coheres with both a decreasing total capital investment (Figure 6-1) and an increasing plant efficiency (Figure 6-2). Capital costs decrease for a large part because of decreasing gas volume in the cleaning section, the extra costs for oxygen compression are more than outweighed by reduced compression costs downstream. The plant efficiency increases because at the higher partial reactant pressure, the selectivity for desired product is higher (at same fixed conversion) even though at atmospheric and air gasification significantly more $CO+H_2$ is produced (Table 3-1), and also because the internal power needs decrease with higher gasification pressure and less nitrogen dilution.

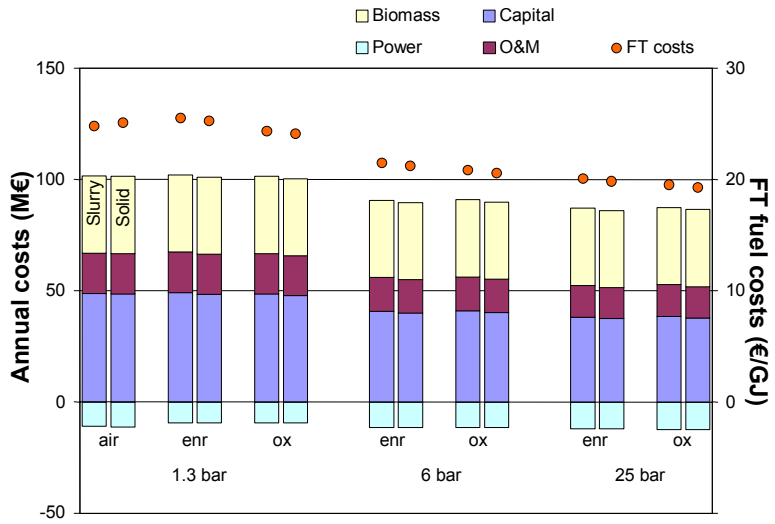

Figure 6-1. Annual costs (bars) and resulting FT liquids production costs (bullets), for different gasification configurations (air, enr and ox stand for air blown, oxygen enriched air blown and oxygen blown). Fixed settings: 400 MW_{th} input scale, tar cracker, wet gas cleaning, no reforming, once through FT, 70 %

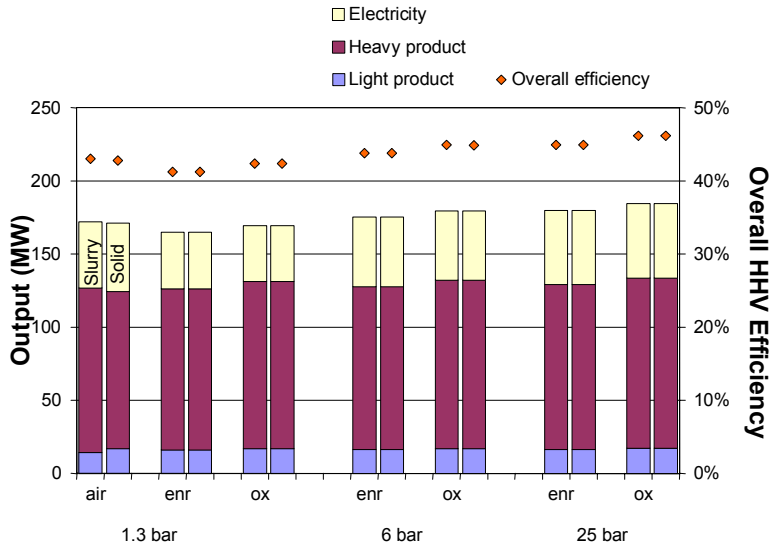

Figure 6-2. Product selectivity (bars) at 400 MW HHV input, and resulting overall plant fuel+electricity efficiency (diamonts) for different gasifier configurations. Fixed settings: tar cracker, wet gas cleaning, no reforming, once through FT, 70 % conversion.

Slurry phase and solid bed reactors perform the same, in the rest of this analysis we only discuss solid bed reactors, as similar results can be expected for slurry phase.

Total Capital Investment for the 400 $MW_{th,HHV}$ input plant (or 168 $MW_{FT,HHV}$ output) is about 286 M€. Without the pre-treatment and gasification section and with a reduced gas cleaning section, the TCI would be 93.3 M€ or 554 €/kW_{FT}. This can be compared with the TCI for grass roots natural gas fed FT plants, reported to amount 300 $MUS\$_{1997}$ for a capacity of 10,000 bbl/day (740 $MW_{FT,HHV}$) or 521 M€/GW_{FT} (Jager 1997), and 750 $MUS\$_{2001}$ for a 2.22 $GW_{FT,\ HHV}$ plant or 394 M€/GW_{FT} (Ghaemmaghami 2001). Taking in account the scale disadvantage, the TCI found thus complies with literature values.

The capital build up for the extremes of these configurations is given in Figure 6-3. The gas cleaning costs, which make out a large part in atmospheric air fired concepts, are strongly reduced in pressurised oxygen fired concepts, where – naturally – the costs for oxygen production and compression play an important role. The shares of costs for pre-treatment, gasifier and power generation grow, but remain equally distributed.

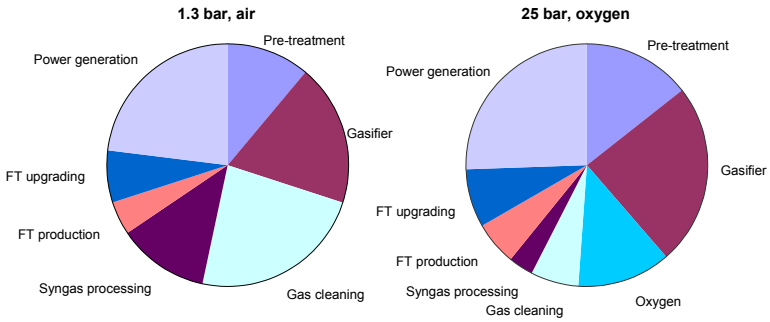

Figure 6-3. Capital break down for two once through concepts: based on air fired atmospheric gasifier (left) and based on oxygen fired 25 bar gasifier (right). Fixed settings: tar cracker, wet gas cleaning, no reforming, solid bed FT, 70 % conversion.

So far, the $CO+H_2$ conversion in the FT reactor was set to 70 %. A higher conversion can be realised by a larger reactor, leading to higher capital costs. Since the capital costs of the FT reactor are not crucial for the total capital investment, a higher conversion leads to lower FT diesel production costs, see Figure 6-4.

According to the graph, the lowest costs are reached at total conversion, this stems from the assumption that reaction rate is constant (the average between entrance and exit reaction speed) over the length of the reactor. Of course, as CO and H_2 disappear over conversion, the reaction rate decreases rapidly

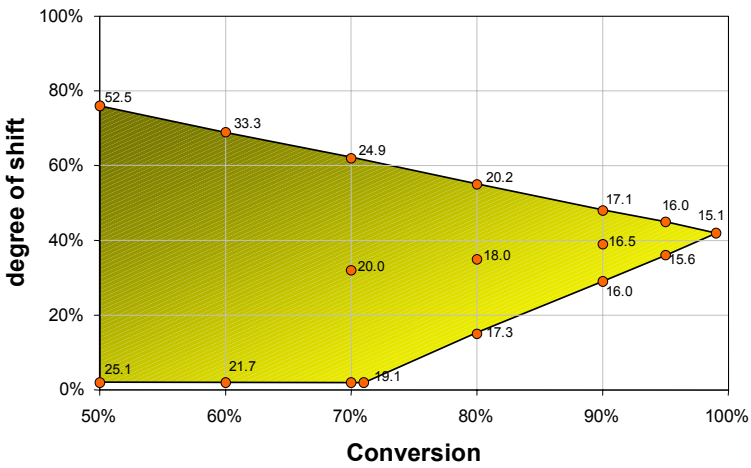

Figure 6-4. FT liquids production costs as a function of conversion extent and degree of shift for 25 bar oxygen gasification (bullets) at 400 MW_{th}. Below area is hydrogen shortage, above area is carbon monoxide shortage. Fixed settings: tar cracker, wet gas cleaning, no reforming, once through FT.

when approaching full conversion, leading to exponential reactor size increase for the last converted reactants. We assume that at 90 % conversion, the entrance-exit-averaged reaction rate is just still representative for the volume-averaged reaction speed. A 400 MWth input system applying conventional technologies can thus produce FT diesel at 16.1 €/GJ, which corresponds to the 14.8 €/GJ (31.2 Dfl_{2000}/GJ) found by Tijmensen for a similar system at the same scale.

The fraction CO_2 before the FT reactor is about 30 volume %, removing this fraction improves both selectivity and efficiency, but due to the accompanying increase in investment this does not result in lower product costs (see Figure 6-5). The efficiency for the dry gas cleaned concepts is slightly higher than wet cleaned concepts, because the ceramic membrane more effectively shifts (since product is taken away) than the traditional shift reactor, hence less steam is needed. This is outweighed by a slightly higher capital investment, such that the resulting fuel price is the same. The oil scrubber effectively returns BTX and tars to the gasifier where they are cracked, but much energy remains in methane, ethane and ethane, so that the FT yield is low. The performance of concepts with oil scrubber improves when adding a reformer, however, this turns out a more expensive combination than a single tar cracker. Overall efficiencies for the best performing systems are 40 – 45 % on HHV basis.

The recycle is operated at 70 % conversion of $CO+H_2$ per pass, 50 % of the off gas is recycled so that overall conversion is about 90 %, like in the once through concepts. The FT pressure is decreased to 25 bar to avoid very large pressure leaps. Nevertheless the continuous temperature leaps and product leaps (syngas partly to C_{1-4} product, and than back to syngas) lead to a both low selectivity and low overall efficiency, and the capital costs are high. The FT 25 bar once through concepts is included to show that the lower operation pressure contributes only marginally to the increased FT liquids production costs.

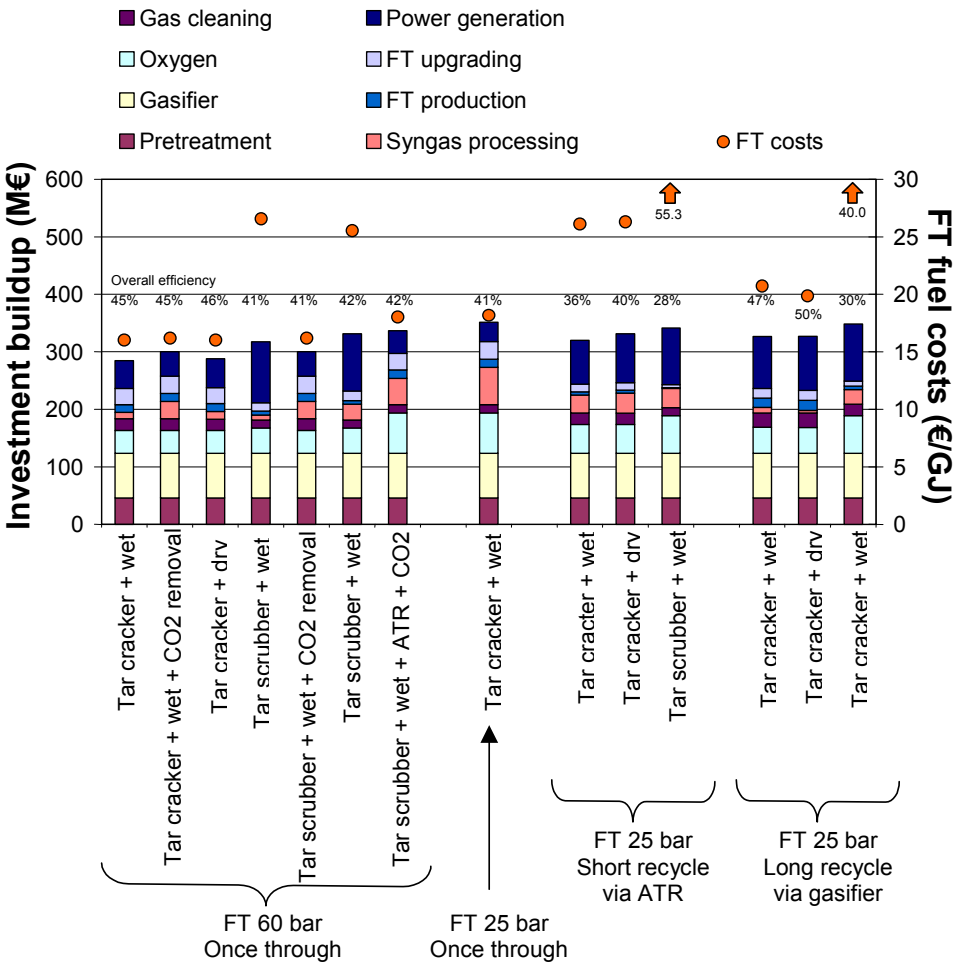

Figure 6-5. Variations on the key concept (25 bar oxygen gasification and 90 % $CO+H_2$ conversion), applying other gas cleaning trains, reforming and/or CO_2 removal, once through or recycle mode. The overall (fuel+electrical) HHV efficiency is indicated.

The influence of scale (see Figure 6-6) is examined for both wet and dry cleaned concepts. It is found that at large scale, FT diesel production costs can drop below 14 €/GJ. The resulting overall scale factor for both concepts is found to be 0.78 (Tijmensen assumed 0.91). The efficiency advantage of dry cleaned concepts does not lead to lower costs at larger scales, compared to wet cleaned concepts.

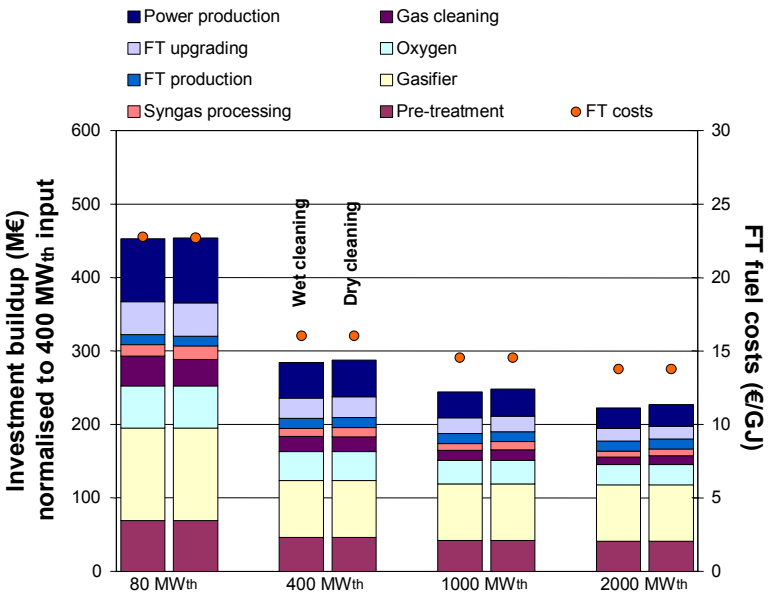

Figure 6-6. Influence of scale on relative investment costs (bars, left axis), and resulting FT liquids production costs (bullets, right axis) for wet and dry cleaned concepts. Fixed: 25 bar oxygen fired gasification, Tar cracker, no reforming, once through 90 % conversion.

The uncertainty of this economic evaluation mainly follows from the ± 30 % uncertainty on capital investment, which directly acts on Operating and Maintenance. Figure 6-7 shows that this results in ± 18 % uncertainty in the FT fuel costs. The biomass price (± 30 % uncertainty) brings ± 13 % uncertainty in the FT fuel cost. The calculated efficiency has a much lower uncertainty (assumed to be ± 5 %) acting on the produced amount of FT fuel and electricity. Its influence on the FT fuel cost is small. Also uncertainties in the electricity price (± 10 %) have little influence on the FT product.

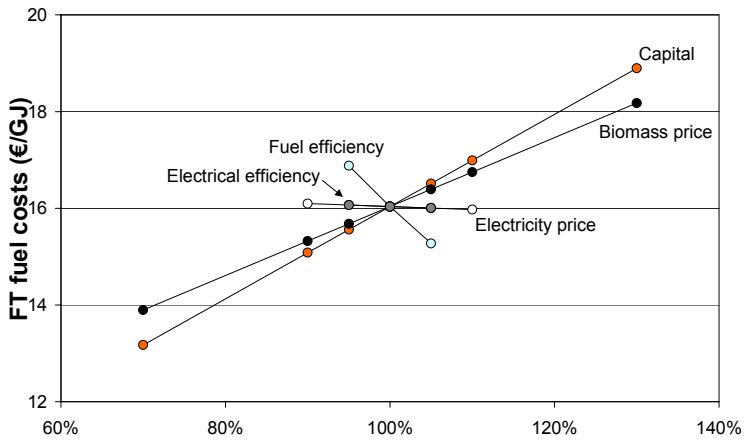

Figure 6-7. Influence of variation of uncertain parameters on FT liquid production costs. Fixed: 400 MW_{th} input system, 25 bar oxygen fired gasification, Tar cracker, wet cleaning, no reforming, once through 90 % conversion.

6.2 Towards the ultimate concept

In the previous paragraph it was shown that, at present, FT diesel could be produced at 16.1 €/GJ by applying conventional technologies, at the reasonable scale of 400 MW_{th} input. This is still about four times the production costs for fossil diesel, which are about 2.6 – 7.0 €/GJ (BP 2002). However, as fuel production costs are only about one-third of the sales price, FT diesel from biomass can become competitive with fossil diesel when the fuel is (partly) exempted from excise duty and VAT (11.6 €/GJ and 3.5 €/GJ, respectively, in the Netherlands).

Looking at the longer term (~15 years), more (cost) improvements are possible. Production at a larger scale may bring the FT fuel costs to below 14 €/GJ (recall Figure 6-6). Furthermore, the biomass price may drop to 2 €/GJ, decreasing the FT production costs to 14 €/GJ (at 400 MW_{th} input) or even 11.5 €/GJ (2000 MW_{th}). Many technical developments may also improve the overall economic system performance, several of these are summarised in Table 6-1. Technological learning could reduce capital costs with 15 % for the third generation plant built, which reduces the FT costs with about 9 %. A major breakthrough would be the development of catalysts that would selectively produce a desired product fraction, *e.g.* the diesel fraction, avoiding the large amount of longer chain hydrocarbons that have to be cracked down again. If this pure 'diesel' selectivity could be realised, the upgrading section could be much cheaper or may even be omitted, implying another 5 – 10 % total capital cost reduction. The joint effect of large scale, technological learning, and selective catalyst brings the FT diesel costs to 9 €/GJ.

Table 6-1. Options that may improve the technical and economical performance.

	Near term	Long term perspective	Impact
Oxygen production		Cryogenic separation more efficient, energy use could decrease from 300 kWh_e/tonne to theoretical minimum of 30 kWh_e/tonne (van Ree 1992)	Limited
		Gas turbine driven membrane separation cheaper[1)], however, availability of this technology is unknown	Significant
Gasification	High carbon conversion or recover energy from char in ash		2-3 %-points on gasification efficiency
Tar removal		Catalytic tar cracker operating at lower temperature: 900 °C instead of 1300 °C	High
		Physical/chemical instead of thermal tar removal technologies	High
FT catalyst	α Slightly higher		Limited
		Selectivity towards desired product C_{10-20}, e.g. by sterical hindered catalysts, thus decreasing upgrading costs	Very high
Gas turbine		General efficiency increases slightly	Limited
Steam cycle		Higher steam pressures	Limited
Process integration	Improve heat integration		Limited
	Find the optimum between energy consumption in biomass drying and gain in gasification step		Limited
	Avoid compressor intercooling by balancing compression before and after gasification		Unknown
Scale		Larger scale decreases capital, labour and may be feedstock costs	Limited
Technological learning		More experience with large scale gasification, logistics, more automation	High

1) Costs may eventually be 0.038 $€_{2002}/m_{NTP}^3$, while cryogenic separation costs about 0.05 $€_{2002}/m_{ntp}^3$ (van Ree 1992).

Green FT diesel at present is only competitive with fossil diesel when produced at a scale of at least 100 – 200 MW_{th} input, and when exempted from taxes. The competitiveness of future large scale FT production strongly depends on the evolvement of an international biomass market, with large quantities of biomass against low prices, and ongoing catalyst development. The technology roadmap

towards such future is quite straightforward as on the longer term the technology will basically be similar to that applied in this study. It can be expected that single devices may improve their efficiency, and there may be continuing competition between dry gas cleaning and wet gas cleaning, solid bed FT reactors and slurry reactors.

Two other biofuels, methanol and hydrogen, were previously studied by Hamelinck and Faaij (2001). Produced at similar scale and biomass price (400 MW_{th}, available technologies, 3 $US\$_{2001}$/GJ) these fuels cost 11.6 and 10.9 €/GJ respectively. On the longer term, at large scale (2000 MW_{th}), with advanced technologies, reckoning with technological learning and at a biomass price of 2 US$/GJ, these costs decrease to 8.3 and 7.5 €/GJ. Although the recalled study cannot directly be compared with the present one, it seems that FT diesel is more expensive both on the short and long term. Besides the energy costs of the different alternative transportation fuels, other factors determine whether FT diesel becomes a success or not, such as its applicability in existing distribution structures, improvements in ICEV (internal combustion engine vehicle) technology, the availability of the FCV (fuel cell vehicle) and realisation of its potentially high efficiency.

7 Conclusion and recommendation

The technological and economic performances of all promising biomass to FT conversion concepts are evaluated using one dynamic Aspen Plus flowsheet. The influence of each parameter or device, on investment costs, FT and electricity efficiency and resulting FT diesel costs is directly visible. Compared with the previous study done by Tijmensen, the present study offers flexibility to explore these influences, the performance of system components is modelled more precise, heat, hydrogen and oxygen are integrated in the flowsheet if applicable, and the economic analysis is done with greater detail.

Total Capital Investment or TCI for a 400 MW_{th} input plant (or 168 $MW_{FT,HHV}$), consisting of a 25 bar oxygen fired gasifier, followed by a tar cracker and wet gas cleaning and a solid bed FT reactor with 70 % once through conversion, is about 286 M€. A high conversion (90% of $CO+H_2$) can be realised in once through mode by application of a large reactor. This results in higher capital costs, but since the capital costs of the FT reactor are not crucial for the total capital investment, higher conversion leads to lower FT diesel production costs. Overall efficiencies for the best performing systems are 40 – 45 % on HHV basis. With such systems FT liquids can be produced at 16 €/GJ.

System variations to the above named concept give the following insights:

- Removing the 30 volume % CO_2 fraction prior to the FT reactor improves both selectivity and efficiency, but due to the accompanying increase in investment this does not result in lower product costs.
- The efficiency for the dry gas cleaned concepts is slightly higher than wet cleaned concepts as less steam is needed. This is outweighed by a slightly higher capital investment, such that the resulting fuel price is the same.
- The oil scrubber effectively returns BTX and tars to the gasifier where they are cracked, but much energy remains in methane, ethane and ethane, resulting in lower FT yields. The performance of concepts with oil scrubber improves when adding a reformer, however, this turns out to be a more expensive combination than a single tar cracker.
- Application of a recycle loop instead of once through does not decrease the production costs: The continuous temperature leaps and product leaps lead to both a low selectivity and a low overall efficiency. Furthermore, the capital costs are high.

The FT liquids production costs for the short term are still about two to four times the production costs for Fossil diesel. Fossil diesel costs strongly depend on the oil price, and could go up. Green diesel could be exempted from excise duty and VAT (11.6 €/GJ and 3.5 €/GJ in the Netherlands) to value the environmental benefits of green FT diesel. The combined effect possibly makes FT diesel from biomass competitive with fossil diesel.

On the longer term (~15 years), more cost improvements are foreseen. The combined effects of larger production scale (2000 MW_{th}) and cheaper biomass (2 €/GJ) may bring the FT fuel production costs to 11.5 €/GJ. The technical and economical performance may be improved by for example developments in oxygen production and gasification, application of catalytic tar cracking instead of thermal, increased selectivity towards the desired product in FT synthesis. These technical developments combined with technological learning (capital costs reduce with 15 % for the third generation plant) bring the FT diesel costs to 9 €/GJ. These values for short and long term are comparable with the costs found by Tijmensen. However, his short term biomass price assumed was 2 €/GJ (vs. 3 €/GJ in present study), and his long term scale was 1600 MW_{th} input (vs 2000 MW_{th}), so that effectively the short term FT costs here, are lower than found by Tijmensen, and the long term costs somewhat higher.

The quality of the analysis depends partly on the model and partly on the economic parameters. The influence of the model is most visible in the Fischer Tropsch reaction, where selectivity and conversion rate are defined as a function of reactant concentration, ratio, pressure and temperature. The relations for selectivity and reaction rate, as presented in Equation 4-2, Equation 4-4 and Table 4-2, are strong simplifications of reality. However, they make the influence of process conditions on performance quite visible and thus add an important improvement compared to previous work in this field. The availability of more public accessible data of recent experiments, would improve the model.

The factored estimation method yields investment costs with an uncertainty range of ± 30 %, resulting in ± 3 €/GJ_{FT}. For several devices, insight in the factors that determine the investments is not available

to the academic world. This makes it difficult to discuss improvement options and (esp. gasifiers) scale up possibilities. The poor availability economic data in general, often leads to repeatedly quoting old quotes, and losing the meaning of the economic numbers.

It is recommended that further research focuses on the further development of oxygen blown pressurised biomass gasification, especially the larger scale applications, and that more insight is acquired in the technical maximum of such gasifiers and the accompanying costs. Furthermore, continuing fundamental catalyst development is necessary to couple a high selectivity (of only the desired product) to a high conversion. And eventually the competitiveness of future large scale FT production strongly depends on the evolvement of an international biomass market, with large quantities of biomass against low prices.

8 Literature

Adcock KD, Fain DE, James DL, Marshall BB, Phillips MR, Powell LE and Raj T, 1999, Ceramic membrane for high temperature gas separation, in proceedings of Thirteenth Annual Conference on Fossil Fuel Energy Materials, Oak Ridge.

Adeyiga AA, 2000, Development of attrition resistant iron-based Fischer-Tropsch catalysts, Technical progress report number 1, Department of Chemical Engineering/Hampton University, Hampton, 11.

Air Liquide, 2002, Website, http://www.airliquide.be/nl/business/products/pipeline/index.asp.

Air Products, 2002, Website information on oxygen http://www.airproducts.com/oxygen/.

Air Products, 2002, Website information on pressure swing adsorption technology http://www.airproducts.com/hydrocarbon/.

Arthur D. Little, 1999, Analysis and evaluation of GAVE chains, GAVE report 9921, Novem, Utrecht the Netherlands, 31 pp.

Bannock G, Baxter RE and Davis E, 1998, The Penguin Dictionary of Economics, online available at http://www.xrefer.com/.

Bartholomew CH, 1990, Recent technolgical developments in Fischer-Tropsch catalysis, Catalysis Letters 7: 303-316.

Bauen A, 2001, Gasification-based biomass fuel cycles: An economic and environmental analysis, in: Kyritsis S, Beenackers AACM, Helm P, Grassi A and Chiaramonti D (eds.), proceedings of First World Conference and Exibition on Biomass for Energy and Industry, James x James, London UK, 143-146.

Bechtel, 1991-1994, Baseline design/economics for advanced Fischer-Tropsch technology, Quarterly Reports 1991-1994, US Department of energy, Pittsburgh.

Bechtel, 1996, Baseline design/economics for advanced Fischer-Tropsch technology, Topical report VI Natural gas Fischer-Tropsch case, Volume II Plant design and Aspen process simulation model, Pittsburgh PA USA, 90 pp + annexes.

Bechtel, 1998, Aspen process flowsheet simulation model of a Battelle biomass-based gasification, Fischer-Tropsch liquefaction and combined-cycle power plant, Bechtel, Pittsburgh, USA, 60 pp + appendices.

Bergman PCA, van Paasen SVB and Boerrigter H, 2002, The novel "OLGA" technology for complete tar removal from biomass producer gas, in proceedings of Pyrolysis and Gasification of Biomass and Waste Expert Meeting, Strasbourg France.

Boerrigter H, den Uil H and Calis H, 2002, Green diesel from biomass via Fischer-Tropsch synthesis: New insights in gas cleaning and process design, in proceedings of Pyrolysis and Gasification of Biomass and Waste Expert Meeting in Strasbourg.

BP, 2002, BP statistical review of world energy, available at www.bp.com/centres/energy/, Bournemouth UK, 44 pp.

Christensen TS and Primdahl II, 1994, Improve syngas production using autothermal reforming, Hydrocarbon Processing 73(3): 39-44.

Consonni S and Larson E, 1994, Biomass-gasifier/aeroderivative gas turbine combined cycles. Part A : Technologies and performance modeling, in proceedings of The American Society of Mechanical Engineers' 8th Congress on Gas Turbines in Cogeneration and Utility, Industrial and Independent Power Generation, Portland.

DeLallo MR, Buchanan TL, Klett MG, Rutkowski MD and White JS, 1998, Decarbonized fuelproduction facility baseline plant - draft letter report, Parsons Power Group, Inc., Reading, Pennsylvania, 14 pp + annexes.

den Uil H, de Smidt RP, van Hardeveld RM, Calis HP, Faaij APC, Hamelinck CN and Tijmensen MJA, 2001, Co-production of liquid fuels and electricity from biomass using a BIG FT CC system: Technical and economic feasibility (confidential report, in Dutch), ECN, Petten, 75 pp + annexes.

Dierendonck LL, 1990, Scale increase, craft or science (in Dutch), PT Procestechniek 1990(2): 20-23.

Dry ME, 1981, The Fischer-Tropsch process, in: Anderson JR and Boudart M (eds.), Catalysis: Science and Technology, Springer, Berlin, Germany, 160-253.

EIA, 2002, International Energy Outlook, online available via www.eia.doe.gov/oiaf/ieo/index.html, Energy Information Administration, Washington DC USA, 273 pp.

Espinoza RL, Steynberg AP, Jager B and Vosloo AC, 1999, Low temperature Fischer-Tropsch synthesis form a Sasol perspective, Applied Catalysis A: General 186: 13-26.

Faaij A, Meuleman B and van Ree R, 1998, Long term perspectives of biomass integrated gasification with combined cycle technology, Netherlands agency for energy and the environment Novem, Utrecht, 93 pp + annexes.

Faaij A, van Ree R and Oudhuis A, 1995, Gasification of biomass wastes and residues for electricity production: Technical, economic and environmental aspects of the BIG/CC option for the Netherlands, Department of Science, Technology and Society/Utrecht University, Utrecht.

Fain DE, 1997, The amazing promise of inorganic membranes, in proceedings of 15th Annual Membrane Technology/Planning Conference, Membrane Technology, Boston.

Geerlings JJC, Wilson JH, Kramer GJ, Kuipers HPCE, Hoek A and Huisman HM, 1999, Fischer-Tropsch technology - from active site to commercial process, Applied Catalysis A: General 186: 27-40.

Ghaemmaghami B, 2001, GTL: Progress and prospects - study yields generic, coastal-based GTL plant, Oil and Gas Journal 99(12).

Greil C, 2002, Personal communication (with Tulleken) on large scale CFB gasification and fuel feeding to boiler, Lurgi AG.

Hamelinck CN, Faaij APC, Larson E and Kreutz T, 2001, Future prospects for the production of methanol and hydrogen from biomass, Science, Technology and Society/Utrecht University, Utrecht, the Netherlands, 28 pp + annexes.

Hendriks C, 1994, Carbon dioxide removal from coal-fired power plants (PhD thesis), Utrecht University/Department of Science, Technology and Society, Utrecht, the Netherlands, 260 pp.

Iglesia E, Reyes SC, Madon RJ and Soled SL, 1993, Selectivity control and catalyst design in the Fischer-Tropsch synthesis: sites, pellets, and reactors, Advances in Catalysis 39: 221-302.

Jager B, 1997, Developments in Fischer-Tropsch technology, Natural Gas Conversion IV - Studies in Surface Science and Catalysis 107: 219-224.

Katofsky RE, 1993, The production of fluid fuels from biomass, Center for Energy and Environmental Studies/Princeton University, Princeton.

Kirschner MJ, 1999, Oxygen, in: Uhlmann's Encyclopedia of Industrial Chemistry, Sixth edition, Electronic release, Wiley-VCH.

Kwant KW, 2001, Status of Gasification in countries participating in the IEA Bioenergy gasification activity, IEA Bioenergy/Novem, Utrecht the Netherlands, 74.

Larson E, Consonni S and Kreutz T, 1998, Preliminary economics of black liquor gasifier/gas turbine cogeneration at pulp and paper mills, in proceedings of The 43rd ASME gas turbine and aeroengine congress, exposition and users symposium, Stockholm.

Larson ED and Marrison CI, 1997, Economic scales for first-generation biomass-gasifier/gas turbine combined cycles fueled from energy plantations, Journal of Engineering for Gas Turbines and Power 119: 285-290.

Liinanki L, Lindmann N, Sjöberg SO and E. S, 1985, Methane yield from biomass gasification at high temperature and pressure, in: Overend RP, Milne TA and Mudge LK (eds.), Fundamentals of thermochemical biomass conversion, Elsevier science Publications Inc., New York, pp923-936.

Livingston EH, 1993, Build you working knowledge of process compressors, Chemical Engineering Progress(2): 27-36.

Louw S, 2002, Personal communication on solving polynomial equations, Utrecht University>Helmholtz Institute, Utrecht the Netherlands.

Maretto C and Krishna R, 1999, Modelling of a bubble colmn slurry reactor for Fischer-Tropsch synthesis, Catalysis Today 52(2-3): 279-289.

Maretto C and Krishna R, 2001, Design and optimisation of a multi-stage bubble column slurry reactor for Fischer-Tropsch synthesis, Catalysis Today 66(2-4): 241-248.

Milne TA, Evans RJ and Abatzoglou N, 1998, Biomass gasifier 'tars': their nature, formation and conversion, report NREL/TP-570-25357, National Renewable Energy Laboratory, Golden, USA, 204.

Mitchell SC, 1997, Hot gas particulate filtration, IEA Coal Research, London, 88 pp.

Mitchell SC, 1998, Hot gas cleanup of sulphur, nitrogen, minor and trace elements., IEA Coal Research, London, 84 pp.

OECD, 1996, National accounts - Main aggregates, Edition 1997, Paris, 170 pp.

Oonk H, Vis J, Worrell E, Faaij A and Bode J-W, 1997, The MethaHydro-proces - preliminary design and cost evaluation, TNO, The Hague.

Parsons I&TG, 1998, Decarbonized fuel plants utilizing inorganic membranes for hydrogen separation, Presentation at 12th Annual Conference on Fossil Energy Materials, Knoxville, 12-14 May, Parsons Infrastructure and Technology Group, Inc.

Perlack RD, Wright LL, Huston MA and Schramm WE, 1995, Biomass Fuel from Woody Crops for Electric Power Generation, report ORNL-6871, available at http://bioenergy.ornl.gov/reports/fuelwood/toc.html, Oak Ridge National Laboratory, Oak Ridge Tennessee USA.

Perry RH, Green DW and Maloney JO, 1987, Perry's chemical engineers' handbook, sixth edition, McGraw-Hill Book Co, Singapore.

Peters MS and Timmerhaus KD, 1980, Plant design and economics for chemical engineers, Third edition, McGraw-Hill Book Company, New York, USA, 973 pp.

Phyllis, A database containing information of the composition of biomass and waste, Netherlands Energy Research Foundation, www.ecn.nl/phyllis/, Petten.

Pierik JTG and Curvers APWM, 1995, Logistics and pretreatment of biomass fuels for gasification and combustion, Netherlands Energy Research Foundation ECN, Petten.

Schulz H, 1999, Short history and present trends of Fischer-Tropsch synthesis, Applied Catalysis A: General 186(1-2): 3-12.

Shell, 2001, Shell SMDS Website, http://www.shell.com.my/smds/.

Sie ST and Krishna R, 1999, Fundamentals and selection of advanced Fischer-Tropsch reactors, Applied Catalysis A: General 186(1-2): 55-70.

Solantausta Y, Bridgewater T and Beckman D, 1996, Electricity production by advanced biomass power systems, Technical Research Centre of Finland (VTT), Finland.

Tijmensen MJA, Faaij APC, van Hardeveld MRM and Hamelinck CN, 2002, Exploration of the possibilities for produciton of Fischer Tropsch liquids via biomass gasification, Biomass and Bioenergy 23(2002): 129-152.

Turkenburg WC, 2000, Chapter 7, Renewable energy technologies, in: Goldemberg Jet al (eds.), World energy assessment report, United Nations Development Programme, New York USA.

van der Laan G and Beenackers T, 2000, Kinetics, selectivity and scale up of the Fischer-Tropsch synthesis, NPT procestechnologie(3): 18-21.

van der Laan GP, 1999, Kinetics, Selectivity and Scale Up of the Fischer-Tropsch Synthesis (Ph.D. Thesis), University of Groningen, Groningen, 238.

van Dijk KM, van Dijk R, van Eekhout VJL, van Hulst H, Schipper W and Stam JH, 1995, Methanol from natural gas - conceptual design & comparison of processes, Delft University of Technology, Delft, 286 pp.

van Ree R, 1992, Air separation technologies - An inventory of technologies for 'pure' oxygen production for pulverised coal combustion in a $CO_2(g)/O_2(g)$-atmosphere (in Dutch), ECN, Petten, 94 pp.

van Ree R, Oudhuis A, Faaij A and Curvers A, 1995, Modelling of a biomass integrated gasifier/combined cycle (BIG/CC) system with the flowsheet simulation programme ASPEN+, Netherlands Energy Research Foundation ECN and Utrecht University/Science, Technology and Society, Petten.

Walas SM, 1987, Rules of thumb, selecting and designing equipment, Chemical Engineering(March 16): 75-81.

Wilhelm DJ, Simbeck DR, Karp AD and Dickenson RL, 2001, Syngas production for gas-to-liquids applications: technologies, issues and outlook, Fuel processing technology 71: 139-148.

Williams RH, Larson ED, Katofsky RE and Chen J, 1995, Methanol and hydrogen from biomassa for transportation, with comparisons to methanol and hydrogen from natural gas and coal, PU/CEES Report 292, Princeton University/Center for Energy and Environmental Studies, Princeton, 47 pp.

Wiltenburg A, 2002, July 9th, Personal communication on cryogenic and adsorbtive oxygen production, Air Liquide, Eindhoven, the Netherlands.

Yates IC and Satterfield CN, 1991, Intrinsic kinetics of the Fischer-Tropsch synthesis on a cobalt catalyst, Energy & Fuels 5: 168-173.

Annex A Oxygen production

Several options exist for the supply of large amounts of oxygen, depending on required amount and purity, see Figure A-1 and Figure A-2. Oxygen could be produced on-site by adsorption or cryogenic distillation technology. Pressure Swing Adsorption (PSA) is the preferred technology for lower purity oxygen (< 93.5 %). PSA has lower investment costs, but a higher energy use, which makes it suitable for maximally 120 – 150 tpd. Above this range – even for purity < 93.5 % – cryogenic air separation is always more economical, the lower purity in that case would be achieved by mixing 96.5 % oxygen with air (Wiltenburg 2002). This means that there is barely any cost advantage when the same amount of oxygen is needed.

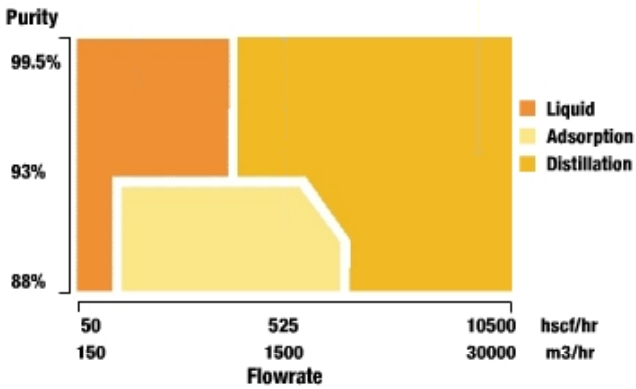

Figure A-1. A map of Air Products' oxygen supply capabilities: Liquid storage of truck delivered oxygen, on-site oxygen production through pressure swing adsorption, or cryogenic distillation (Air Products 2002).

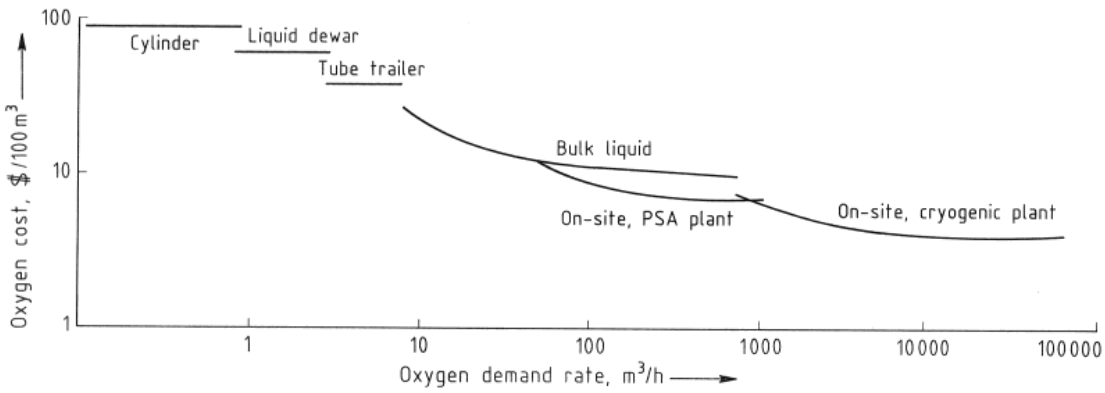

Figure A-2. Cost of oxygen as a function of rate of use and supply mode (Kirschner 1999 in Uhlmann's Encyclopedia of Industrial Chemistry).

Investment cost data from several other studies are summarised in Table A-1 and Figure A-3.

Table A-1. Investment costs data for cryogenic air separation units from other studies in M€$_{2002}$.

	Base costs (M€)	Base scale (tpd)	R	installation const.	var[1)]	max scale (tpd)[2)]	Purity	Remarks
Van Dijk et al (1995)	39.9	1000	0.8508	0 %	0 %	2000	99.5 %	27 C, 1.2 bar
Williams et al (1995)	41.05	1000	0.712	45 %	25 %	1000	95 %	-
Williams et al (1995)	47.22[3)]	1000	0.712	45 %	25 %	1000	99.5 %	-
Larson et al (1998)	26.88	1100	0.6	0 %	0 %	1100	unknown	135 C, 27 bar
Bechtel (1996)[4)]	42.06	2037	0.7	73 %	0 %	2045	99.5 %	27 C, 12 bar
Tijmensen (2002)[5)]	26.6	576	0.75	30 %		576	95 %	-
Tijmensen (2002)	27.9[6)]	576	0.75	30 %		576	99.5 %	-

1) Part of the installation costs are fixed and part decrease with scale, see Footnote under Table F-2.
2) Highest maximum scale found so far is 3200 tonne/day O_2 (Shell Bintulu Wilhelm et al. 2001).
3) Capital costs increase with 15 % to produce 99.5 % purity instead of 95 %.
4) Here base costs are ISBL, the installation factor includes offsites (OSBL), and HO Service and contingency. The source does not mention working capital or start up costs.
5) Here base costs are "installed costs" (ISBL + engineering, CNH), the installation factor includes building interest (5%), contingency (10 %) and start-up costs (5%). OSBL costs and working capital are not mentioned by the source.
6) Capital costs increase with 5 % to produce 99.5 % purity instead of 95 %.

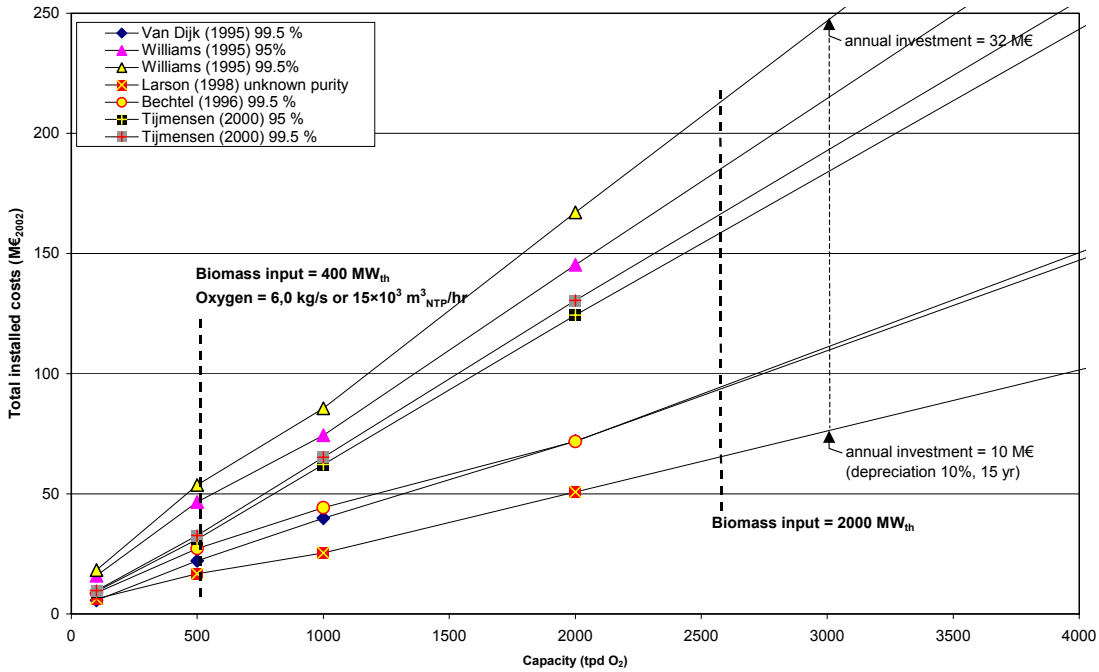

Figure A-3. Cost lines for cryogenic air separation units in five different studies.

Clearly in the Table "base costs" plus "installation" do not necessarily resemble the total capital investment, or TCI (see Annex F). Therefore these studies give a large range of total costs. Uhlmann's Encyclopedia of Industrial Chemistry is an authoritative source of information. The 0.050 €/m$^3_{NTP}$ oxygen for a 3000 tpd supply (753·10^6 m$^3_{NTP}$/yr) in Figure A-2 imply that the annual costs are 38 M€, which includes the annual capital depreciation (13 % of total installed costs), O&M (4 %), and energy costs (300 kWh/tonne or about 8.6 M€ annual), so that the total installed costs will have been about 173 M€. Therefore it is assumed that the cost lines by Tijmensen best represents the total capital investment.

Between Northern France and Rotterdam (Netherlands) and at the US Gulf Coast, a unique oxygen distribution pipeline is available (Air Liquide), when high purity oxygen (99.5% and up) is required (see). The pipeline ensures continuous and flexible oxygen delivery, and may offer cost advantages (Wiltenburg 2002).

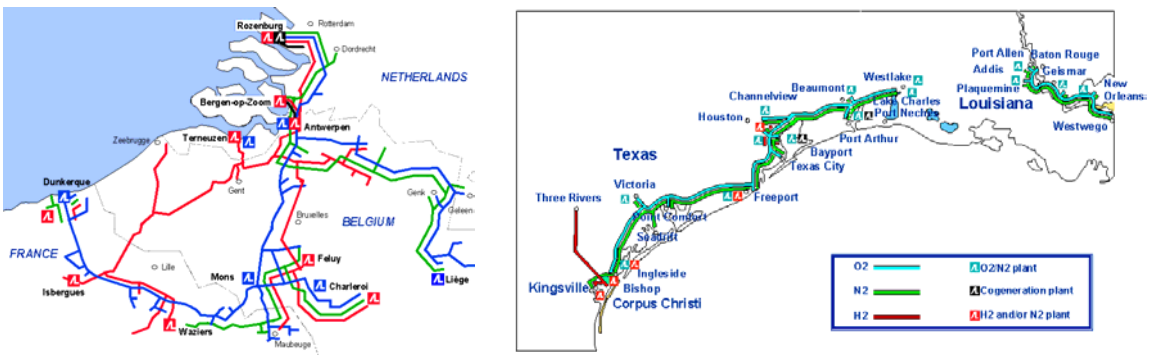

Figure A-4. Air Liquide pipeline networks for industrial gases in Belgium (left) and US Gulf Coast. Large air separation units and oxygen pipelines are indicated in blue (Air Liquide 2002).

Compression of oxygen requires a non-lubricated compressor, because oil and grease in contact with oxygen, especially at high pressures, can ignite explosively (Air Products 2002; Wiltenburg 2002). This kind of compressor costs 1.4 times as much as a lubricated compressor.

Annex B FT synthesis

B.1 FT theory

Hydrocarbon liquid synthesis over iron or cobalt catalysts was invented by Fischer and Tropsch in the mid 1920s and went through a lively history since (Sie and Krishna 1999; Schulz 1999; Dry 1981). In wartime Germany and isolated South Africa, strategic arguments for liquid fuel production from coal exceeded economic aspects. In the 1970s FT regained attention as a response to fading oil reserves and oil boycotts. The actual interest in FT synthesis stems from environmental demands, technological developments and changes in fossil energy reserves, e.g. the necessary conversion of remote gas into shippable liquids.

B.1.1 α the chain growth probability, and S_{C5+} the liquid selectivity

The Fischer Tropsch synthesis produces chains of different length and can be modelled as a chain growth reaction of CO and hydrogen on the surface of a cobalt or iron catalyst:

$$CO \quad + \quad 2H_2 \quad \rightarrow \quad -CH_2- \quad + \quad H_2O$$ Equation B-1

This reaction is exothermal, about 20% of the chemical energy is converted to heat, and gives a big decrease in molar volume. Therefore the reaction is carried out intermediate temperature and pressure (225 – 365 °C, 5 – 40 bar). Heat has to be removed to maintain optimum catalyst life and reaction rate.

Chain termination takes predominantly place as an olefin (Tijmensen et al. 2002). The variety of chain lengths is determined by the probability that a hydrocarbon chain propagates or terminates, and is described by the Anderson-Schulz-Flory (ASF) distribution. This simple statistical model predicts a linear relationship between the logarithm of the molar amount of a paraffin and its carbon number. The chain growth probability is expressed by the constant α. The molar yield of a hydrocarbon with n carbon molecules is:

$$C_n = \alpha^{n-1}(1-\alpha)$$ Equation B-2

with α = chain growth probability
n = length of hydrocarbon
C_n = fraction of hydrocarbons having length n

The product distribution is shown in Figure B-1. Modern Fischer-Tropsch processes trend towards higher selectivities: α approaches 1. The product with a high average molecule mass, is to be converted to gasoline, Diesel and other products in the product recovery and upgrading section. Then Diesel fuel selectivities approaching 80 % can be obtained (Schulz 1999).

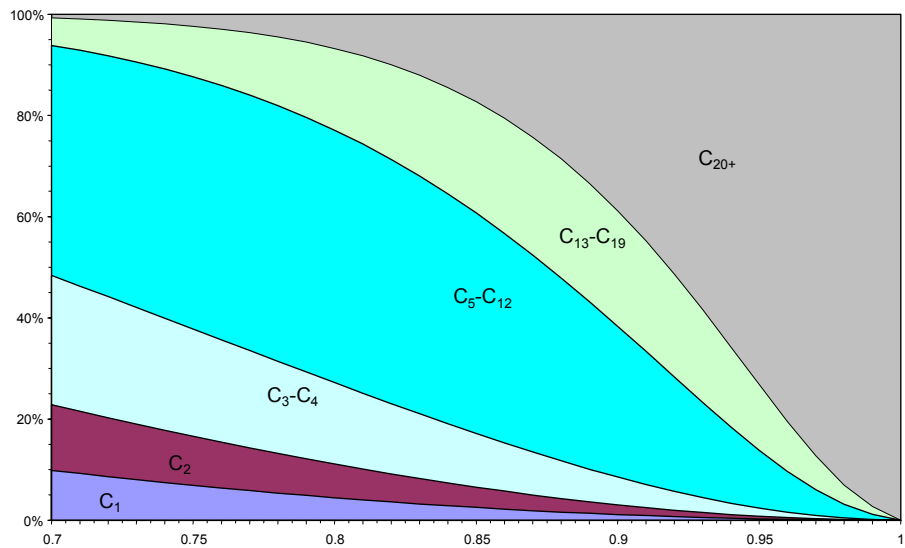

Figure B-1. FT product distribution for different α, calculated using Equation 2-3.

In practice, product distribution differs significantly from the theoretical ASF distribution (Geerlings et al. 1999; van der Laan and Beenackers 2000), this is mainly due to hydrogenolysis and reinsertion of olefins at a growth site:

$$C_n + H_2 \rightarrow C_{n-1} + CH_4$$ Equation B-3

This causes an increase in C_1 and a decrease in C_{5+} selectivity. Other deviations from the ASF distribution are a lower than expected C_2 and an increased amount of C_3 and C_4. Also, at very high molecule mass selectivity may slightly increase (Tijmensen et al. 2002).

The ASF distribution is named as an inherent limitation for premium products (olefins, gasoline, Diesel) selectivity (Bartholomew 1990), since a wide range of products below and above is co-produced. Most attempts to circumvent the selectivity limitations have met with failure. Nevertheless, there are new approaches involving shape selective supports, unsteady-state operation, and interception of intermediates which show promise for improving selectivity beyond that predicted by ASF theory (Bartholomew 1990).

Water, formed as primary product can react with CO to form CO_2 via the water gas shift reaction:

$$H_2O + CO \Leftrightarrow CO_2 + H_2$$ Equation B-4

Generally speaking, the shift activity of cobalt catalysts is negligible (Maretto and Krishna 1999).

The reaction takes place in a three-phase system: gas (CO, H_2, steam, and gaseous hydrocarbon products), liquid product and solid catalyst. The amounts of syngas and product molecules that have to be transferred between the phases are quite large: i.e. an order of magnitude larger than the amount of hydrogen molecules to be transferred in hydroprocessing of oils. Therefore, great demands are placed on the effectiveness of interfacial mass transfer in FT synthesis (Sie and Krishna 1999).

B.1.2 Influencing selectivity

The growth probability or selectivity, is largely catalyst dependent (Shell 2001) and depends furthermore on temperature, local partial pressures of reactants and inerts, and the applied FT technology (Dry 1981; Schulz 1999; Tijmensen et al. 2002). For example, α increases with decreasing H_2/CO ratio, decreasing reaction temperature and increasing pressure, and α is higher for Ru and Co catalysts relative to Fe catalysts (Bartholomew 1990).

Reactants

The FT synthesis consumes H_2 and CO at a ratio depending on selectivity. With $\alpha = 0$ the H_2/CO consumption ratio is 3, with $\alpha = 1$ the ratio is 2. The other way around is also true: a lower H_2/CO ratio in the reacting gas increases the selectivity by decreasing the termination rate (Tijmensen et al. 2002). Therefore the consumption ratio will be higher than the actual H_2/CO ratio, hence the H_2/CO ratio shifts over the reactor length, and the selectivity will be different at different locations in the reactor, or intermediate hydrogen injection is necessary to maintain the actual ratio. Additionally decreasing H_2/CO ratio decreases the catalyst activity and productivity (Tijmensen et al. 2002; den Uil et al. 2001)

The H_2/CO ratio may be adjusted during the FT reaction by a water gas shift occurring on the catalyst surface, or by reactant injection at different places in the reactor. Iron catalysts strongly enhance the shift reaction, cobalt catalysts only slightly (see below). Shift during FT synthesis can be especially useful to increase the conversion extent, while maintaining a relative hydrogen shortage for high selectivity. It may be necessary to inject additional water/steam in the reactor. If the shift reaction in the FT reactor is too fast, shortage of CO will decrease the selectivity and restrict the conversion.

A too high H_2O/H_2 and CO_2/CO ratio oxidises the Fe catalyst; it then undergoes attrition. This can cause fouling/plugging of downstream filters and equipment, makes the separation of catalyst from the oil/wax product very difficult if not impossible, and results a steady loss of catalyst from the reactor (Adeyiga 2000). On the other hand too much CO can lead to coking effects on the catalyst. To eliminate these problems the reaction extent is limited by a shortage of CO.

The presence of inert does not directly influence the selectivity, only by decreasing the partial pressure of reactants (see below). It is assumed that all inert components (CO_2, CH_4, N_2, Ar and light

hydrocarbons) behave the same with respect to the FT reaction. With shift enhancing catalysts, CO_2 is a reaction product, instead of an inert. Low concentrations of CO_2 are not expected to influence selectivity or conversion and are considered inert. CO_2 can even be a reactant, if H_2:CO > R as is the case with syngas from the IGT+ gasifier.

Temperature
The actual selectivity is very different from that expected from thermodynamic calculations (Dry 1981). The influence of temperature on the selectivity is consistent for all F-T catalysts: High temperature reduces the average chain length of the product molecules, favours selective methane formation, and favours deposition of carbon and thereby deactivation of (particularly the iron) catalyst. On the other hand high temperature favours a high reaction rate. As the FT process is strongly exothermic, it is essential to keep isothermal process conditions to maintain a high reaction rate without the adversities of influencing the reaction product. Therefore temperature control is a major subject in reactor choice later (Schulz 1999).(Tijmensen et al. 2002).

The temperature, depending on the operation mode, ranges from 180 to 250 °C for the *low temperature FT*, or from 300 to 350 °C for the *high temperature FT*.

(Partial) pressure
The FT process is generally operated at pressures ranging from 20 to 40 bar. Higher reactant pressure leads to higher selectivity.

Catalyst type
Cobalt, iron, nickel and ruthenium catalyse the FT synthesis. Nickel, however, reacts with CO at elevated pressures and its liquid selectivity is not appealing. Ruthenium is of high scientific interest, being most active, working at the lowest reaction temperature (as low as 150 °C), and producing the highest molecular weight hydrocarbons. Unfortunately its high price and limited world resources exclude industrial application (Schulz 1999).

Iron based catalysts are cheap and exhibit a high selectivity to olefins, oxygenates, and light hydrocarbons (Maretto and Krishna 1999). They have water gas shift activity, a favourable feature for FT synthesis with CO-rich syngas, as usually is obtained from biomass gasification (Schulz 1999). The catalyst can thus be operated at significantly lower H_2/CO ratios, e.g. 0.6 – 1.0, compared to values of 2 for Co and Ru catalysts, without significant deactivation due to carbon deposition by the Boudouard reaction (Bartholomew 1990). On the other hand Maretto and Krishna (1999) state that iron catalysts deactivate rapidly due to coke deposition on catalyst surface and oxidative reactions. The activity of iron catalysts decreases through product inhibition by water. This feature restricts the conversion degree and necessitates a gas recycle operation (Schulz 1999).

The high density of iron sites per unit surface area will result in a higher intrinsic activity. But with advancing conversion and increasing water partial pressure, this advantage over supported cobalt catalysts disappears (Espinoza et al. 1999).

Cobalt catalysts have a negligible water gas shift activity (Maretto and Krishna 1999). They generate higher molecular weight hydrocarbons (paraffinic waxes), promote hydrogenation and produce limited amounts of oxygenates (Maretto and Krishna 1999). A sufficiently high CO partial pressure – even at the catalyst bed end – should be maintained in order to avoid excessive methane formation (Schulz 1999). Today cobalt catalysts for FT Diesel production from natural gas are designed for a maximum wax selectivity (Schulz 1999). With cobalt catalysts olefin readsorption on FT sites takes place and this contributes significantly to the desired high wax selectivity (Schulz 1999). Cobalt catalysts have an FT reaction rate more related to the ratio of the partial pressures of hydrogen/carbon monoxide, compared to iron catalysts whose FT reaction rate is more related to the absolute partial pressures of the reactants. Cobalt catalysts are thus more productive at lower space velocities and higher water partial pressures, whereas iron catalysts are even inhibited by the produced water. These conducts allow cobalt to achieve higher per pass conversions than iron (Espinoza et al. 1999). The higher conversions with cobalt lead to a reduction of about 30 % in the capital required for FT section of the plant (Jager 1997). The main disadvantage is that cobalt allows less flexibility with respect to the process conditions (Jager 1997).

Cobalt catalysts are much more expensive than iron ones. This necessitates the significantly longer operating life for the cobalt based catalyst (degradation is now practically eliminated) compared to the iron based catalyst (Espinoza et al. 1999; Maretto and Krishna 1999; Schulz 1999). Furthermore cobalt based FT catalysts have advantages for use in slurry reactors, compared to iron based catalysts (Espinoza et al. 1999). No commercial slurry reactor is yet operating using cobalt based catalyst (only iron so far), but the technology for preparation of cobalt based catalysts for slurry reactors is already in place and results of tests and demonstration projects are promising (Espinoza et al. 1999).

FT catalysts can lose activity as a result of oxidation, sintering, fouling and poisoning. Oxidation is only relevant for iron and then only under abnormal conditions: with high CO_2 and H_2O partial pressures. Water vapour clearly enhances the sintering rate of iron catalysts. During production the pores of all catalysts will become wholly or partially filled with wax, which eventually slows down the reaction rate. This calls for catalyst regeneration by solvent washing or cracking with hydrogen. At high pressure, the catalyst activity decrease is slower, but also more difficult to treat. Both iron and cobalt catalysts are poisoned by sulphur compounds as H_2S, C_2H_5SH, COS and thiophene; the sulphur content in the synthesis gas should be below 0.2 mg S/m_{NTP}^3 (0.02 ppm) for iron catalysts (Dry 1981). As both extensive cleaning and catalyst replacement have certain costs, there will be an optimum. For cobalt catalysts, the maximum amount of sulphur in the feed is expected to be much lower.

Table B-1. Influence of the sulphur content of synthesis gas on the rate of activity decline for fluidised iron catalyst at about 593 K (Dry 1981).

Sulphur in syngas mg S/m_{NTP}^3	Daily drop in percent conversion	Catalyst half life (days)
0.1	very low	-
0.4	0.25	277
2.8	2.0	34
28	33	1.7

Under the FT synthesis conditions normally used for nickel, cobalt and ruthenium based catalysts little deposition of carbon occurs. Carbon monoxide deposits on iron at temperatures above 550 K by the Boudouard reaction, resulting in particle swell, disintegration and ultimately plugging of the reactor (Dry 1981).

B.2 Reactor type

There are three main kinds of FT reactors: the fixed bed reactor, the fluidised bed reactor and the slurry phase reactor. Of all these reactor types commercial versions exist (Shell, Sasol, Exxon). Comparisons between the reactor types are broadly found (Bartholomew 1990; Sie and Krishna 1999; Tijmensen et al. 2002).

Fixed bed reactors are suited for low temperature FT synthesis (200 – 240 °C), aiming at a high average molecular weight of the product (Schulz 1999). In fixed bed reactors, the particle size (> 1 mm) and its corresponding intraparticle diffusion can be a limiting factor for the overall reaction rate and also increases the pressure drop. This large particle size is necessary to control heat removal. Conversion of more than 70 % is possible, but to obtain an even temperature profile, conversion is usually limited to 20-30 % per pass and the gas is recycled (Sie and Krishna 1999). The multi tubular fixed bed reactor is for example used in the Shell Middle Distillate synthesis (SMDS) process.

Since the FT reaction is highly exothermic, gas-solid fluidised beds with their excellent heat transfer and temperature equalisation characteristics seem very attractive. Fluidised beds also have the advantage that small catalyst particles, e.g. of about 100 μm, can be used. These are free from diffusion limitation, contrary to the larger particles in fixed beds. However, a serious issue is the possibility that heavy product deposits on the catalyst, causing particles to agglomerate and thus hampering fluidisation. This can be overcome by operating above the hydrocarbon dew point, i.e. at high temperature, low pressure, and low conversion. However, this in turn yields a low α (< 0.71) product, and rules out the possibility of applying gas-solid fluidised beds for FT processes that produce much heavier products than gasoline (Sie and Krishna 1999). Bartholomew (1990) states that fluidised beds lead in system throughput per unit volume of reactor, but this is not confirmed by others since.

Also the slurry reactor is very well mixed and can thus operate isothermally. The absence of axial and radial temperature gradients – present in fixed bed reactors – allows much higher average operating

temperatures in the slurry reactor and hence higher reaction rates (Sie and Krishna 1999; Jager 1997). Due to the small size of catalyst particles (50 μm) intraparticle diffusion is not a limiting factor. For high reactor productivity, the slurry reactor needs to be operated at high superficial syngas velocity and highest handleable catalyst concentrations (up to 40 vol %). In the heterogeneous regime the hold-up decreases and the bubbles are larger (1 to 10 cm compared to millimetres in the homogeneous regime), but due to the continuous break-up and coalescence, the effective bubble diameter remains small and the high gas-liquid mass transfer remains high (Sie and Krishna 1999). Reducing the overall backmixing of the slurry phase, by staging, increases the syngas conversion per pass (above 0.95 is possible) and the reactor productivity per unit of volume (Maretto and Krishna 2001). The isothermal higher average temperature operation together with the more effective contact between the syngas and the catalyst particles leads to higher production rates for the same overall reactor dimensions even though the catalyst inventory in the slurry reactor is lower. Because all the catalyst is at more optimal process conditions, not only are the yields per reactor volume higher, but the catalyst consumption per ton of product is only 20 to 30 % of that of the fixed bed reactor (Jager 1997). Catalyst loading and unloading in slurry reactors is much easier, activity can even be upheld by replacement during run (Sie and Krishna 1999).

A slurry type reactor is much simpler in construction than the TFB reactor. The suspended cooling coils and a gas distributor give a much cheaper arrangement than the tube and tube sheet arrangement in the fixed bed reactor. Because of this the slurry reactor lends itself much better to scale up. At high gas throughputs, the fixed bed reactor becomes very complex and mechanically difficult to manufacture and the pressure drops across the tubes becomes excessive. The pressure drop across the fixed bed reactor varies from 3 – 7 bar, depending on operating pressure, whereas it is less than 1 bar across the slurry reactor (Jager 1997). Operating costs are much reduced (Jager 1997)

The maximum feasible capacity of the different reactors is not fixed. If the weight would be limited to 900 tonnes per reactor (for transportation) the multi-tubular fixed bed reactor has a capacity of 500 tonne/day, and the slurry reactor 1250 to 2500 tonne/day (= 20000 bbl/day) (Sie and Krishna 1999). At present operating pressures of 25 bar, single slurry reactors with capacities of 10.000 bbl/day can be designed. Still higher capacities at higher pressures are possible. The cost of a single 10.000 bbl/day reactor train is about 25 % of that of an equivalent fixed bed system with six reactors (Jager 1997).

Overall, the slurry reactor appears to be the most efficient, economical system for production of light olefins and gasoline and the best match for syngas form advanced gasifiers (Bartholomew 1990). The slurry concept is regarded the most efficient process for FT clean diesel production (Schulz 1999).

Table B-2. Factors determining reactor choice.

	Multi-tubular fixed bed	**Fluidised bed**	**Slurry Bubble column**
Construction	Complex[1)]	-	Simple[1)]
Heat exchange	limited conversion per pass and large particle size necessary to obtain even temperature profile[2)]	Excellent heat transfer results in isothermal conditions[2)]	Excellent heat transfer results in isothermal conditions[3)]
Solids separation	not needed[2)]	Filtering may be necessary[2)]	Filtering is necessary[2)]
Gas-liquid separation	-	Foam formation occurs[2)]	-
Reactant distribution	Difficult[2)]	Automatically lateral mixed[2)]	Automatically lateral mixed[2)]
Catalyst settling or agglomeration	-	-	Too low velocities may lead to concentration gradient, insoluble materials may deposit on catalysts and hamper proper suspension[2)]
Catalyst deactivation	The top region (where the syngas enters) acts as a sulphur trap, whereas in the bottom region almost no sulphur is found but water inhibition plays a more important role.	-	Sulphur poisoning and water inhibition have evenly effect over the reactor (continuous circulation), leading to 1.5 – 2 times higher conversion loss than in fixed bed. High costs of cobalt catalysts necessitates a more

			effective sulphur removal[4)]
Loading/unloading catalyst	difficult, shutdown necessary[2)]	on-line[2)]	on-line[1,2)]
Scale-up	relatively straight forward by multiplying tubes[2)]	More complex and more costly, empirical scale-up through demonstration stage[2)]	Complex, but rational scale-up possible[2)]
Maximum capacity	-	-	2.5 – 6 times larger than fixed bed (if reactor size is weight limited)[1,2)] 2500 tonne/day (about 20,000 bbl/day) may be achieved[2)]

1) Jager (1997).
2) Sie and Krishna (1999).
3) Van der Laan (2000).
4) Espinoza (1999).

FT synthesis cannot achieve in one step greater than 40 – 50 % selectivity for a given premium product such as light olefins, gasoline. However, it is possible through a combination of two or three reaction/separation stages to achieve selectivities on the order of 80 – 90 % for almost any desired product (Bartholomew 1990).

Although in absolute terms the cost savings in the Fischer-Tropsch reactor section may be substantial, their impact on the total plant economy may be modest (Sie and Krishna 1999). To achieve an optimum in performance for the complete process, the catalyst and the reactor should be optimised in a combined fashion (Geerlings et al. 1999).

Annex C Mathematic relations

C.1 Reactant consumption ratio

The ratio in which CO and H_2 are consumed depends on the selectivity.

$$CO + 3H_2 \rightarrow CH_4 + H_2O$$ Equation C-1

$$2CO + 5H_2 \rightarrow C_2H_6 + 2H_2O$$ Equation C-2

$$3CO + 7H_2 \rightarrow C_3H_8 + 3H_2O$$ Equation C-3

$$4CO + 9H_2 \rightarrow C_4H_{10} + 4H_2O$$ Equation C-4

$$5CO + 11H_2 \rightarrow C_5H_{12} + 5H_2O$$ Equation C-5

etc.

Short chains need relatively more hydrogen (H_2:CO closer to 3) than long chains (H_2:CO approaches 2). The ratio between reactant consumption is expressed by weighting the reactant consumption in each of the above reactions by the selectivity towards that reaction:

$$\frac{-R_{H_2}}{-R_{CO}} = \frac{3\cdot(1-\alpha)+5\cdot\alpha\cdot(1-\alpha)+7\cdot\alpha^2\cdot(1-\alpha)+9\cdot\alpha^3\cdot(1-\alpha)+\cdots}{1\cdot(1-\alpha)+2\cdot\alpha\cdot(1-\alpha)+3\cdot\alpha^2\cdot(1-\alpha)+4\cdot\alpha^3\cdot(1-\alpha)+\cdots}$$ Equation C-6

Or:

$$\frac{-R_{H_2}}{-R_{CO}} = \frac{3+5\cdot\alpha+7\cdot\alpha^2+9\cdot\alpha^3+\cdots}{1+2\cdot\alpha+3\cdot\alpha^2+4\cdot\alpha^3+\cdots} = \frac{\sum_1^\infty (2n+1)\cdot\alpha^{n-1}}{\sum_1^\infty n\cdot\alpha^{n-1}}$$ Equation C-7

$$\frac{-R_{H_2}}{-R_{CO}} = \frac{\sum_1^\infty 2n\cdot\alpha^{n-1}+\sum_1^\infty \alpha^{n-1}}{\sum_1^\infty n\cdot\alpha^{n-1}} = 2+\frac{\sum_1^\infty \alpha^{n-1}}{\sum_1^\infty n\cdot\alpha^{n-1}} = 2+\frac{\sum_0^\infty \alpha^{n}}{\sum_0^\infty (n+1)\cdot\alpha^{n}}$$ Equation C-8

The power series are equal to:

$$\sum_0^\infty \alpha^n = \frac{1}{1-\alpha}$$ Equation C-9

$$\sum_0^\infty (n+1)\cdot\alpha^n = \sum_0^\infty n\cdot\alpha^n + \sum_0^\infty \alpha^n$$ Equation C-10

$$\sum_0^\infty n\cdot\alpha^n = 0+\alpha+2\cdot\alpha^2+3\cdot\alpha^3+4\cdot\alpha^4+\cdots = \alpha\cdot(1+2\cdot\alpha+3\cdot\alpha^2+4\cdot\alpha^3+\cdots)$$ Equation C-11

$$\sum_{0}^{\infty} n \cdot \alpha^n = \alpha \cdot (1+\alpha+\alpha^2+\cdots)\cdot(1+\alpha+\alpha^2+\cdots) = \alpha \cdot \left(\sum_{0}^{\infty} \alpha^n\right)^2$$ Equation C-12

$$\sum_{0}^{\infty} (n+1)\cdot \alpha^n = \alpha \cdot \left(\frac{1}{1-\alpha}\right)^2 + \frac{1}{1-\alpha} = \frac{\alpha}{(1-\alpha)^2} + \frac{1-\alpha}{(1-\alpha)^2} = \frac{1}{(1-\alpha)^2}$$ Equation C-13

Thus

$$\frac{-R_{H_2}}{-R_{CO}} = 2 + \frac{\frac{1}{1-\alpha}}{\frac{1}{(1-\alpha)^2}} = 3-\alpha$$ Equation C-14

C.2 Relation between α and C_{5+} selectivity

First in molar fractions.

$$\alpha_{Cn} = \alpha^{n-1}\cdot(1-\alpha)$$ Equation C-15

$$\alpha_{C5+} = 1-\alpha_{C1}-\alpha_{C2}-\alpha_{C3}-\alpha_{C4}$$ Equation C-16

$$\alpha_{C5+} = 1-\alpha^0\cdot(1-\alpha)-\alpha^1\cdot(1-\alpha)-\alpha^2\cdot(1-\alpha)-\alpha^3\cdot(1-\alpha)$$ Equation C-17

$$\alpha_{C5+} = 1-1+\alpha-\alpha+\alpha^2-\alpha^2+\alpha^3-\alpha^3+\alpha^4 = \alpha^4$$ Equation C-18

Similarly it can be deduced that selectivity towards all chains longer than a C atoms is expressed by:

$$\alpha_{Ca+} = \alpha^{a-1}$$ Equation C-19

The C_{5+} selectivity is often expressed in mass fraction S_{C5+}. A molar fraction f is expressed by a mass fraction S_f:

$$S_f = \frac{\overline{M}_f \cdot f}{\overline{M}_{total}}$$ Equation C-20

with $\overline{M}_f$ = average molar mass of the fraction f
$\overline{M}_{total}$ = average molar mass of the total mixture

Using the average molecular mass for C_{a+}, as deduced in Section C.4.

$$S_{Ca+} = \frac{M_{Ca+}\cdot\alpha_{Ca+}}{M_{C1+}}$$ Equation C-21

$$S_{Ca+} = \frac{\frac{14\cdot\alpha+(14a+2)\cdot(1-\alpha)}{(1-\alpha)}\cdot\alpha^{a-1}}{\frac{14\cdot\alpha+16-16\cdot\alpha}{(1-\alpha)}} = \frac{(14\cdot\alpha+(14a+2)\cdot(1-\alpha))\cdot\alpha^{a-1}}{16-2\cdot\alpha}$$ Equation C-22

$$S_{C5+} = \frac{(14\cdot\alpha+72\cdot(1-\alpha))\cdot\alpha^4}{16-2\cdot\alpha} = \frac{72-58\cdot\alpha}{16-2\cdot\alpha}\cdot\alpha^4 = \frac{36-29\cdot\alpha}{8-\alpha}\cdot\alpha^4$$ Equation C-23

It is difficult – if not impossible – to express α exactly as function of S_{C5+}. However, Equation F-23 can be approached by F-24 (Louw 2002). The fit is shown in Figure C-1.

$$S_{C5+} = \frac{1}{e^{\left(-\frac{(1-\alpha)^2}{0.1392}\right)}}$$ Equation C-24

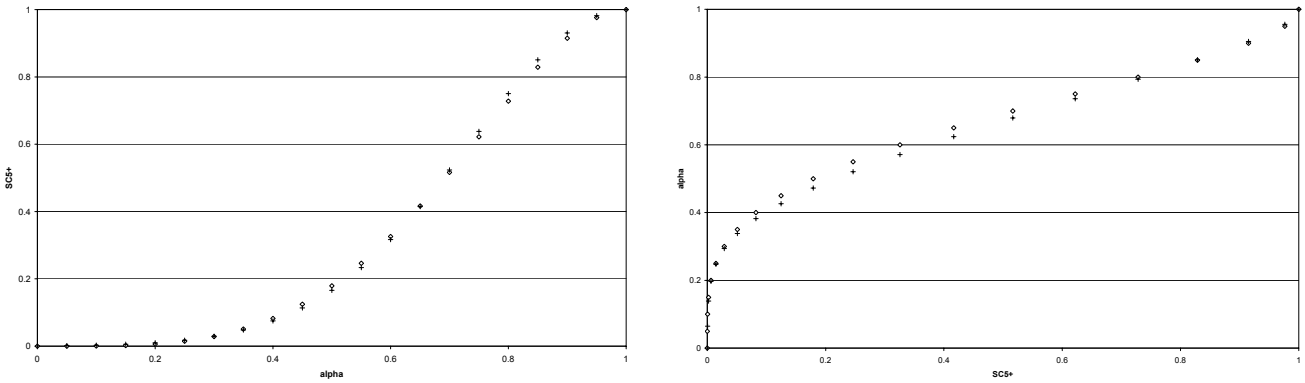

Figure C-1. S_{C5+} as function of α (left), and vice versa (right). Diamonds express the exact relation of Equation F-23, crosses express the fit of Equation F-24 and F-25.

The approached expression for α becomes:

$$\alpha = 0.75 - 0.373 \cdot \sqrt{-\log(S_{C5+})} + 0.25 \cdot S_{C5+}$$ Equation C-25

Via similar ways follows any weight selectivity from α:

$$S_{Ca} = \frac{M_{Ca} \cdot \alpha_{Ca}}{\overline{M}_{total}} = \frac{(14a+2)\cdot(\alpha^{a-1}\cdot(1-\alpha))}{\frac{16-2\alpha}{1-\alpha}} = \frac{(7a+1)\cdot(\alpha^{a-1}-2\alpha+\alpha^{a+1})}{8-\alpha}$$ Equation C-26

C.3 Hydrogen deficiency in hydrocracking

All C_{20+} chains are hydrocracked to a length of preferably C_{10} and maximal C_{19}:

$$C_{20}H_{42} + H_2 \rightarrow 2C_{10}H_{22}$$ Equation C-27

$$C_{25}H_{52} + H_2 \rightarrow C_{10}H_{22} + C_{15}H_{32}$$ Equation C-28

$$C_{48}H_{98} + 3H_2 \rightarrow 3C_{10}H_{22} + C_{18}H_{38}$$ Equation C-29

The number of hydrogen atoms needed in the hydrocracking process to cut a chain into smaller parts is called the hydrogen deficiency. $C_{20\text{-}29}$ chains have a deficient of 1 H_2, $C_{30\text{-}39}$ chains have a deficient of 2 H_2, etc. Using Equation F-19:

$$\alpha_{C10-19} = \alpha^9 - \alpha^{19}$$ Equation C-30

The average hydrogen efficiency can thus be expressed by:

$$D_{H_2} = \frac{\sum_0^\infty n \cdot \left(\alpha^{10n+9} - \alpha^{10n+19}\right)}{\alpha^9} = \frac{\sum_0^\infty n \cdot \alpha^9 \cdot \alpha^{10n} - \sum_0^\infty n \cdot \alpha^{19} \cdot \alpha^{10n}}{\alpha^9}$$ Equation C-31

$$D_{H_2} = \frac{\left(\alpha^9 - \alpha^{19}\right) \cdot \sum_0^\infty n \cdot \alpha^{10n}}{\alpha^9} = \left(1 - \alpha^{10}\right) \cdot \sum_0^\infty n \cdot \alpha^{10n}$$ Equation C-32

$$D_{H_2} = (1-s) \cdot \sum_0^\infty n \cdot s^n = \frac{(1-s) \cdot s}{(1-s)^2} = \frac{\alpha^{10} - \alpha^{20}}{\left(1-\alpha^{10}\right)^2}$$ Equation C-33

D_{H2} is average deficiency per chain.

C.4 Average molecular mass

For all chains with n C atoms holds the following selectivity and molecular mass:

$$\alpha_{Cn} = \alpha^{n-1}(1-\alpha)$$ Equation C-34

$$M_{Cn} = 12n + 2n + 2$$ Equation C-35

All chains that have *a* C atoms and longer, therefore, have average molecular mass:

$$M_{Ca+} = \frac{\sum_a^\infty (14n+2) \cdot (1-\alpha) \cdot \alpha^{n-1}}{\alpha^{a-1}} = \frac{(1-\alpha) \cdot \sum_a^\infty \left(14n \cdot \alpha^{n-1} + 2 \cdot \alpha^{n-1}\right)}{\alpha^{a-1}}$$ Equation C-36

$$M_{Ca+} = \frac{(1-\alpha) \cdot \sum_0^\infty \left(14 \cdot (n+a) \cdot \alpha^{n+a-1} + 2 \cdot \alpha^{n+a-1}\right)}{\alpha^{a-1}}$$ Equation C-37

$$M_{Ca+} = \frac{(1-\alpha) \cdot \alpha^{a-1} \cdot \sum_0^\infty \left(14n \cdot \alpha^n + 14a \cdot \alpha^n + 2 \cdot \alpha^n\right)}{\alpha^{a-1}}$$ Equation C-38

$$M_{Ca+} = (1-\alpha) \cdot \left(14 \cdot \frac{\alpha}{(1-\alpha)^2} + (14a+2) \cdot \frac{1}{1-\alpha}\right) = \frac{14 \cdot \alpha + (14a+2) \cdot (1-\alpha)}{(1-\alpha)}$$ Equation C-39

Annex D Unit modelling assumptions

Table D-1. Unit modelling assumptions.

General	
Heat exchanger[1,2,3)]	$\Delta p/p$ = 2%; $T_{HOT} - T_{COLD}$ > 10K
Compressor[4)]	$\eta_{Isentropic}$ = 0.78; η_{mech} = 1 Compression ratio is same for each stage, maximum is 4, such that outlet temperature does not exceed 250°C; Intercooling to 25 °C – 130°C, last stage no duty Oxygen compressor: 1 → 6 bar: 7.2 kW_e; 1 → 25 bar: 16.8 kW_e
Pretreatment	
Steam dryer[5)]	Specific heat consumption is 2800 MJ/twe, or 0.95 kg steam at 12 bar 235 °C (extracted from steam cycle) per kg moist evaporated specific electricity consumption is 65 kWh_e/twe
Cryogenic Oxygen production[6)]	
	Electricity consumption is 0.3 MWh_e/tonne O_2
Gasification	
	Gasification at 1.3, 6, or 25 bar, with oxygen, enriched air, or air. Composition of produced gas in Table 4-1. Gasification of recycle flows and scrubbed tar is modelled as partial oxidation with gasification medium at gasification pressure (Gibbs minimisation reactor), equivalence ratio is set φ = 0.3
Gas cleaning section	
Cyclones	Remove ash and carbon prior to tar cracker/tar scrubber. $\Delta p/p$ = 2%
Tar cracker	Partially combusts the fuel gas with pure oxygen to supply heat for cracking reaction. Cracker is fed by same medium as gasifier, such that the exit temperature is raised to 1300 °C. Modelled as two subsequent reactors that are adiabatic connected (2% heat loss): 1) Formation of char (C) 0.5 % of carbon in feed by Calculator TARCRC; 2) Gibbs equilibrium of all components except char.
Tar scrubber	Tars are scrubbed from the gas by oil, and recycled to the gasifier. First the gas stream is cooled to 400 °C with heat exchange for steam raising. Then all components C_6H_6 and $C_{14}H_{10}$ are splitted from the main stream at 100 °C by means of the ECN oil scrubber. Cooling energy is transferred to the oil, and is considered lost. Oil loss is 0.003 kg/kg syngas and ends up in the gasifier. $\Delta p/p$ = 2 %
Quench scrubber[2,7)]	Modelled as Two Outlet Flash drum 1 m^3 water per 1000 m^3 gas Q = 0 W $\Delta p/p$ = 3%
Hot Gas Cleaning[3,8)]	T_{in} = 400 °C $\Delta p/p$ = 15 %
Particle filter	$\Delta p/p$ = 2 %
Guard beds	$\Delta p/p$ = 1%
Gas processing	
Autothermal Reformer[9)]	ATR1 provides heat ATR2 requires T_{in} = 550°C adjust ratio ATR1/ATR2 to T_{out} = 1000°C Overall 2 mole steam injected per mole carbon; some gas streams do not require additional steam for reforming ATR1: Stoichiometric Reactor: complete combustion of all components, using stoichiometric amount of same medium as in gasifier. T = 1000°C. Δp = -0.5 bar ATR2: Gibbs free energy minimisation Reactor; 2% of CH_4 is inert; Ar and N_2 are inert; C_2H_4 and C_2H_6 react completely
Shift Reactor[1)]	steam = 3 * CO – H_2O Part of stream splits to SHIFT reactor such that ratio (H_2 – CO_2) / (CO + CO_2) = 2.05 ± 0.02 after downstream Selexol Modelled as Gibbs free energy minimisation Reactor T_{in} = 330 °C T approach = +10°C Q = 0 W Δp = -0.5 bar Inerts: CH_4, C_2H_4, C_2H_6, Ar, N_2 Steam injected is 3 * CO - H_2O
PSA system[1,10)]	system operating at 14-28 bar, 40°C recycling 80% PSA-A Δp = -0.35 bar 100% CO_2 + H_2O adsorption desorption at 1.3 bar PSA-B Δp = -0.35 bar adsorption of all gas but 84% of H_2 desorption at 1.3 bar

Ceramic Membrane[1, 11)]	steam = CO-H_2O system operating at elevated pressure: 20 bar or higher $T_{in} = T_{out} = 800$ °C Catalytic molecular sieve: shift all CO on surface to H_2, therefore H_2O:CO => 1 at entrance, transport 95 % of H_2 and 0% of others to product stream Product at 1.2 bar Δp depleted stream = -0.1 bar
Selexol[1, 12)]	98% of CO_2 and 100% of H_2O separation $T_{in} = 127$ °C $\Delta p = -0.5$ bar CO_2 released at 1.5 bar
Chemical Reactors	
FT reactor	T = 230 °C, $\Delta p = -2$ bar Product flows follow from Calculator FT Reaction heat raises LP 6 bar steam for steam cycle
Product upgrading	
Distillation	Light component = C_9H_{20}; recovery in distillate = 99% Heavy component = $C_{10}H_{22}$; recovery in distillate = 1 % Stages = 15 "Partial condensor with all vapor" or "Total condensor" Boiler heat (at ~ 250 °C) extracted from steam cycle: 46.15 tph steam of 34.5 bar/368 °C → 25 bar/324 °C per MW boiler heat
HydroCracking[13)]	T = 490 °C cracking: $C_{20}H_{42} + H_2 \rightarrow 2C_{10}H_{22}$ and $C_{32}H_{66} + 2H_2 \rightarrow 2C_{10}H_{22} + C_{12}H_{26}$ Hydrogen supplied by shift+PSA or CM Reaction heat supplied by steam cycle
Power generation	
Advanced Gas Turbine[14)]	Compressor $p_{fuel} = p_{air} = 30$ bar $\eta_{isentropic} = 0.91$ $\eta_{mech} = 0.99$ Combustor modelled as Gibbs minimisation reactor Identify reaction products: H2O, CO2, O2, N2, AR $\Delta p = 0$ bar, Q = 0 W T after turbine expander = 550 ± 2 °C by adjusting Air If this 550 °C can not be achieved by GT alone, duct burning is applied with 2 % excess air Expander: p = 1.2 bar $\eta_{isentropic} = 0.89$ $\eta_{mech} = 0.99$ T after heat exchanger = 100°C
HRSG[15)]	Gas $T_{out} = 100$ °C Water $T_{in} = 15$ °C
Steam raising	With heat at temperatures >550 °C cooled to >100 °C water 88 bar/15 °C is turned to steam 86.2 bar/510 °C 1.022 tph steam per MW heat transferred With heat at temperatures >200 °C cooled to >100 °C water 6.2 bar/15 °C is turned to steam 6 bar/169 °C 1.232 tph steam per MW heat transferred Heat at temperatures lower than 100 °C is not used in the steam cycle
Steam Turbine	Steam of preferably 86.2 bar/510 °C is expanded to 0.04 bar/29 °C Intermediate steam extraction and injection is possible at 61 bar/455 °C; 34.5 bar/373 °C; 25 bar / 331 °C; 12 bar/246 °C; 6 bar/175 °C; 1.3 bar / 107 °C $\eta_{isentropic} = 0.89$; $\eta_{mech} = 0.99$
Composition oxidation medium in volume %	
Air	$O_2 = 20.75$ % $H_2O = 1.01$ % $CO_2 = 0.03$ % $N_2 = 77.29$ % Ar = 0.92 %
$Oxygen_{99.5\%}$	$O_2 = 99.5$ % $N_2 = 0.5$ %
Enriched	$O_2 = 80.0$ % composed from 79.8 % $oxygen_{95\%}$ and 20.2 % air
$Oxygen_{95\%}$	$O_2 = 95$ % $N_2 = 5$ %

1) Katofsky (1993).
2) Consonni and Larson (1994).
3) Tijmensen (2002).
4) Walas (1987)
5) Stora dryer, Pierik and Curvers (1995)
6) The energy use of oxygen production is taken 300 kWh per tonne O_2 (99.5 %) (van Dijk et al. 1995; van Ree 1992).
7) Perry et al. (1987)

8) Only pressure drop is relevant here, since no contaminants are modelled.

9) Autothermal reformer operates at 20 – 70 bar, 850 – 1100 °C, steam to carbon ratio ranges from 0.5 to 3.5 (Christensen and Primdahl 1994). Oxygen is set stoichiometric for oxidation part of ATR. Equilibrium towards syngas is better at lower pressures.

10) Air products PSA technology can selectively separate CO and/or H2 from a product gas (Air Products 2002).

11) Ceramic membranes modelled as hydrogen separation device or HSD (developed by Oak Ridge National Laboratory). Operation at high temperature, surface has shifting capabilities (Fain 1997; Adcock et al. 1999; DeLallo et al. 1998; Parsons I&TG 1998).

12) Actually, half of the CO_2 would be released at 1 bar and half at 4 bar. The net energy demand of a 436 tonne of CO_2 per hour unit amounts 9 MW_e (Hendriks 1994).

13) The hydrogen efficiency of cracking is high and no extra hydrogen is needed.

14) Van Ree et al. (1995).

15) HRSG after GT or boiler. The flue gas can be cooled down to 100 °C without corrosion problems, since the gas is expected to contain less than 100 ppb sulphur (van Ree et al. 1995).

Annex E Flowsheeting options in Aspen Plus

Table E-1. Calculators, all are Excel workbooks

Name	Description	User interaction
AIRGT2	Calculates and writes the stoichiometric amount of air needed in the Gas turbine, only when duct burning is applied to raise temperature before HRSG.	Reveal when duct burning necessary. Combines with DesignSpec TGT2.
FT	Calculates selectivity and reaction speed from temperature, pressure, CO+H2 concentration and H2/CO ratio. Writes product distribution to FT reactor. Writes reactor size to PERFORM.	Choose solid or slurry type. Set conversion or set reactor dimension.
GASIFR-1	Writes gasifier syngas flow from database to flowsheet. Writes amount of oxygen 95 % or 99.5 % to PERFORM. Writes amount of steam to STEAMCYC. Writes medium type and steam/ox ratio to GASIFR-2.	Choose gasifier medium and pressure. Set scale (default 400 MW_{th}).
GASIFR-2	Regulates amount of oxygen and steam for gasification of recycle flow and scrubbed tar.	Set equivalence ratio (default = 0.3).
HYDROGEN	Calculates necessary hydrogen for hydrocracking. Writes necessary steam for WGS. Directs gas flow.	Choose hydrogen separation technology Shift+PSA or Ceramic Membrane.
OXATR	Regulates amount of oxygen for Autothermal reformer.	-
PERFORM	Reads flow and component dimensions from flowsheet. Calculates economic performance. Writes to summary table.	
PRESHEAT	Regulates pressure and heat losses, and necessary compression.	-
RECYCLE	Direct flows throughout plant by setting splitters.	Choose between No recycle, Short recycle, and Long recycle.
STEAMCYC	Writes plant steam supply and demand to steam cycle.	-
WSCRUBBR	Calculates water flow size to scrubber.	-

Table E-2. Design Specifications.

Name	Description	User interaction
OXTARCR1	T exit Tar cracker 1 approaches 1300 °C by adjusting oxygen in	Hide or reveal depending on cleaning train chosen.
OXTARCR2	T exit Tar cracker 2 approaches 1300 °C by adjusting oxygen in	“
CHTARCR1	5 % of Carbon in fuel becomes elemental C (char)	“
CHTARCR2	5 % of Carbon in fuel becomes elemental C (char)	“
HX402	T hydrocracker approaches 490 °C by adjusting T feed stream.	-
TATR	T exit Autothermal reformer approaches 1000 °C by adjusting oxygen in	Hide or reveal depending on ATR use
TGT1	T before HRSG approaches 550 °C by adjusting amount of air to GT	Hide when duct burning necessary
TGT2	T before HRSG approaches 550 °C by adjusting duct	Reveal when duct burning necessary. Combines with Calculator AIRGT2

Annex F Estimating capital investment

Many methods for estimating capital investment are used in the chemical process industry, methods that differ in detail level and accuracy and that differ in approach; a valuable overview is given by Peters and Timmerhaus (1980). In the present study we apply the *Study estimate* or *Factored estimate,* based on knowledge of major items of equipment; it has a probable accuracy range up to ± 30 % (Peters and Timmerhaus). This method was also applied in previous work of Hamelinck (2001) and will not be explained in detail here. Hamelinck (2001) presented an overview Table with system component costs. In the present study a similar Table is used with extensively reviewed data relevant for FT production, Table F-2.

The total installed capital costs for a certain unit consist of various direct and indirect costs as shown in Table F-1. Equipment costs found in literature are often ill defined, and may be anywhere between f.o.b. and total installed capital. Also the applied percentages for additional costs differ greatly between studies. Of course these percentages depend on the specific location: OSBL costs in industrial area may be lower since the necessary infrastructure is readily available, a Middle East location will differ greatly from a Western Europe location. All the factors to calculate TCI from purchased equipment are very dependent on scale, as ISBL relatively decreases with scale because of increasing process integration and OSBL decreases because certain offsites have a fixed same scale for both small and large plants. However, it is never explained how. Peters and Timmerhaus remark that factors should only be used when more accurate data are not available. Knowing this, it is tried in Table F-2 to strictly define the *base costs* and the *overall installation factor* for each unit separately.

Table F-1. Ratio factors for estimation of Total capital investment costs based on f.o.b., purchased equipment, or installed equipment.

			Peters and Timmerhaus (1980)		**Faaij** (1998)[1)]	**Williams** (1995)
			ordinary chemical plant	solid-fluid processing plant		
f.o.b.[2)]						
shipping and insurance						
Direct costs (DC)	ISBL[3)]	Purchased Equipment[4)] (PE)				
		Installation	25 – 55 % of PE	39 % of PE		
		Installed Equipment (IE)[5)]				
		Instrumentation and controls	6 – 30 % of PE	13 % of PE	5 % of IE	
		Piping, installed	10 – 80 % of PE	31 % of PE	4 % of IE	
		Electrical, installed	10 – 40 % of PE	10 % of PE	7 % of IE	
		Buildings, process and auxiliary	10 – 70 % of PE	29 % of PE	1.5 % of IE	
		Civil works			10 % of IE	
	OSBL[6)]	Connection to grid / utilities			5 % of IE	
		Yard improvements[7)]	} 40 – 100 % of PE	10 % of PE	0.5 % of IE	
		Service facilities		55 % of PE		
		Land (if purchase is required)	4 – 8 % of PE	6 % of PE		
		utilities / auxiliaries				25 % of IE
subtotal = Direct costs (DC)						
		Land				1 % of DC
Indirect costs (IC)		Engineering and supervision	5 – 30 % of DC	32 % of PE	15 % of DC	
		Construction expense	6 – 30 % of DC	34 % of PE		
		Contractor's fee		18 % of PE		
		Contingency	5 – 15 % of FCI	36 % of PE	10 % of DC	20 % of DC
		Building interest			10 % of DC	
		Fees / overheads / profits			10 % of DC	10 % of DC
subtotal = Indirect costs (IC)						
Fixed capital investment (FCI)			DC + IC			
Working Capital (WC)			10 – 20 % of TCI	74 % of PE		10 % of DC
Start-up (SU)					5 % of DC	5 % of DC
Total Capital Investment (TCI)			FCI+WC	DC + IC + WC	DC + IC +SU	DC + IC + WC + SU
Which is			**266 – 1136 % of PE**	**487 % of PE**	**200 % of IE**	**183 % of IE**

1) The given factors hold for a 29 MW_e output BIG/CC plant and the direct costs depend on scale: Instrumentation and control 0.3, piping 0.7, electrical systems (or electronic equipment) 0.3, buildings 0.65, connection to grid 0.2, civil works 0.65 (Faaij et al. 1995).

2) *Free on board*: the price of a traded good after loading onto a ship but before shipping, thus *not* including transportation, insurance, and other costs needed to get a good from one country to another (Bannock et al. 1998); cost of equipment ready for shipment from supplier.

3) or *onsite*.

4) Also called *purchase equipment delivered*, *delivered equipment costs*, or *c.i.f.: cost, insurance, and freight*, the price of a traded good including transport cost, meaning that a price includes the various costs, such as transportation and insurance, needed to get a good from one country to another (Bannock et al. 1998).

5) or *installed hardware*.

6) or *offsites*.

7) or *site preparation*.

Table F-2. Costs of system components in $M€_{2002}$ [1].

Unit	Base cost	Scale Factor	Base Scale	Overall installation factor [25]	Maximum Size [26]
Pre-treatment [2]					
Conveyers [3]	0.41	0.8	33.5 wet tonne/hr	2.0	110
grinding [3]	0.48	0.6	33.5 wet tonne/hr	2.0	110
storage [3]	1.16	0.65	33.5 wet tonne/hr	2.0	110
dryer [3]	8.5	0.8	33.5 wet tonne/hr	2.0	110
iron removal [3]	0.43	0.7	33.5 wet tonne/hr	2.0	110
feeding system [3,4]	0.48	1	33.5 wet tonne/hr	2.0	110
Gasification System					
CFB gasifier [5]	44.3	0.7	68.8 dry tonne/hr	1.69	75
Air separation unit 99.5 % O_2 [6]	27.9	0.75	576 tonne/day O_2	1.3	3200
Oxygen compressor [7]	18.1	0.85	13.2 MW_e	1.86	-
Gas Cleaning					
Tar cracker [3]	3.6	0.7	34.2 m^3/s gas	2.0	52
Oil scrubber [8]	1.64	0.7	14.7 m^3_{NTP}/s gas	1	-
Cyclones [3]	3.0	0.7	34.2 m^3/s gas	2.0	180
High-temperature heat exchanger [9]	8.1	0.6	138.1 MW_{th}	1.84	-
Particle filters [3]	1.9	0.65	12.1 m^3/s gas	2.0	64
Scrubbers [3]	3.0	0.7	12.1 m^3/s gas	2.0	64
Dry gas cleaning [10]	35.8	1.0	74.1 m^3/s gas	1.86	-
Guard beds (ZnO + active C) [11]	0.024	1.0	8.0 m^3_{NTP}/s gas	3	
Syngas Processing					
Compressor [12]	12.9	0.85	13.2 MW_e	1.86	-
Autothermal reformer [13]	31.1	0.6	100 m^3_{NTP}/s	2.3	-
Shift reactor [14]	12.2	0.65	8819 kmol CO+H_2/hr	1.81	-
PSA units A+B [15]	32.6	0.7	9600 kmol feed/hr	1.69	-
Ceramic membrane [16]	25.2	0.8	17 tonne H_2/hr	1	-
Selexol CO_2 removal [17]	63.0	0.7	9909 kmol CO_2/hr	1	-
Fischer-Tropsch Production					
Solid bed gas phase FT 60 bar [18]	25.3	1	208 m^3	1.30	-
Slurry phase FT 60 bar [19]	36.5	0.72	362 m^3	1	365
Product upgrading [20]	233	0.7	286 m^3_{FT}/hr	1	
Power Isle [21]					
Gas turbine + HRSG [3,22]	22.0	0.7	26.3 MW_e	2.0	-
Steam turbine + steam system [3,23]	5.9	0.7	10.3 MW_e	2.0	-
Expansion turbine [24]	5.0	0.7	10.3 MW_e	2.0	-

1) Annual GDP deflation up to 1994 is determined from OECD (1996) numbers. Average annual GDP deflation after 1994 is assumed to be 2.5 % for the US, 3.0 % for the EU. Cost numbers of Dutch origin are assumed to be dependent on the EU market, therefore EU GDP deflators are used. 1 $€_{2002}$ = 0.88 $US\$_{2002}$.

2) Total pre-treatment approximately sums up to a base cost of 11.46 $€_{2002}$ at a base scale of 33.5 tonne wet/hour with an R factor of 0.79.

3) Based on first generation BIG/CC installations. Faaij et al. (1995) evaluated a 29 MW_e BIG/CC installation (input 9.30 kg dry wood/s, produces 10.55 m_{NTP}^3 fuel gas/s) using vendor quotes. When a range is given, the higher values are used (Faaij et al. 1998). The scale factors stem from Faaij et al. (1998).

4) Two double screw feeders with rotary valves (Faaij et al. 1995).

5) Direct costs are 29.74 $MUS\$_{1991}$ for a 1650 dry tonne/day input IGT gasifier, R = 0.7 (Williams et al. 1995). Maximum input is 400 MW_{th} HHV (Tijmensen et al. 2002).

6) Various cost numbers with a large distribution are found in literature (van Dijk et al. 1995; Williams et al. 1995; Larson et al. 1998; Bechtel 1996; Tijmensen et al. 2002), see Annex A. Here the cost number from Tijmensen (2002) is used because it bests fits the real oxygen cost price (Kirschner 1999): Direct costs are 27.9 $M€_{2002}$ for a base scale of 576 tpd oxygen of 99.5 % purity, 1 bar, the overall installation factor is constant 30 %, R = 0.75. It is assumed that oxygen of 95 % purity is 5 % cheaper; oxygen of 80 % purity for the scale considered here, will always be mixed from 95% O_2 and air, and thus barely has cost advantages. The highest maximum scale found to date is 3200 tpd (Shell Bintulu, see Wilhelm et al. 2001)

7) Because of explosion risk, oxygen is compressed in a non-lubricated compressor, which is 1.4 times more expensive than lubricated compressors (Livingston 1993).

8) OLGA technology (Boerrigter et al. 2002). Cost numbers are assumed by the authors.

9) High temperature heat exchangers following the gasifier and (in some concepts) at other locations are modelled as HRSG's, raising steam of 86 bar/510 °C. A 39.2 kg steam/s unit (i.e. transferring 138.1 MW_{th}) costs 6.33 $MUS\$_{1997}$ f.o.b., overall installation factor is 1.84 (Larson et al. 1998), which is assumed to consist of 22 % added direct costs (vary with scale) and 50 % added indirect costs (constant), see note 25.

10) Tijmensen (2002) assumes the f.o.b. price for Hot Gas Cleaning equipment to be 30 $MUS\$_{2000}$ for a 400 MW_{th} HHV input. This equals 74.1 m^3/s from a BCL gasifier (863 °C, 1.2 bar). There is no effect of scaling. The installation factor is taken from Faaij et al. (1998) but for 400 MW_{th} input scale (decreased added direct costs), see footnote 25.

11) For 427 MW_{th} LHV biomass input concepts, Tijmensen (2002) assumes that 1 wt% of the fuel S enters the ZnO bed, and two guard beds of 3 m^3 are required. These beds consist of 2300 kg steel, at a steel price 20 Dfl_{2000}/kg, so that each guard bed will cost 50.000 Dfl_{2000} f.o.b., or 150.000 Dfl_{2000} installed. Tijmensens 70 MW_{th} LHV (13 dry tonne/hour biomass, assumed gas yield 100 kmol/dry tonne) corresponds to a local gas flow of 8.0 m^3_{NTP}/s.

12) Katofsky (1993) assumes compressors to cost 700 $US\$_{1993}$ per required kW_{mech}, with an installation factor of 2.1. The relation used in the present study stems from the compressor manufacturer Sulzer quoted by Tijmensen (2002). At the indicated base-scale, total installed costs are about 15 % higher than assumed by Katofsky. Multiple compressors, for fuel gas, recycle streams, or hydrogen, are considered as separate units. The installation factor is taken from Faaij et al. (1998) but for 400 MW_{th} input scale (decreased added direct costs), see note 25.

13) Hamelinck (2001) based on Katofsky (1993) assumes 4.7 $MUS\$_{2001}$ with an overall installation factor of 2.3 for an 1390 kmol total/hour (8.8 m^3_{NTP}/s) unit, R=0.6, where Bechtel (1991-1994) gives a total capital investment of 20.52 $MUS\$_{1992}$ for an ATR unit of 364.7 MMscfd (120 m^3_{NTP}/s) gas feed, which is almost twice as costly. Here we assume the average between the two: 31.1 M€ total capital investment for a 100 m^3_{NTP}/s unit, R=0.6. The installation factor is assumed to consist of variable direct costs (53 % in base case) and fixed indirect costs (50 %), see note 25.

14) Cost numbers from Williams (1995) are used: 9.02 $MU\$_{1995}$ for an 8819 kmol $CO+H_2$/hr reactor, overall installation factor is 1.81 and R = 0.65. Hendriks (1994) writes that a reactor of 350.000 m^3_{NTP}/hr total gas ($CO+H_2$ is 93.3 %, this gives 14.6 Mmol/hr $CO+H_2$) costs 30 $MUS\$_{1994}$ installed (no R given) which corresponds with the line from Williams. Hendriks also gives an annual (sulphur tolerant) catalyst use of 0.7 $MUS\$_{1994}$, but considers this to be part of the operational costs. Numbers from Tijmensen (2002) are considered too low: 0.95 $MDfl_{2000}$ for direct costs plus engineering (add 30 % to obtain total capital investment) for a 2400 kmol/hr unit, R = 0.6.

15) PSA units (excluding the recycle compressor) cost 23 $MUS\$_{1993}$ for a 9600 kmol feed/hour throughput, R=0.7 (Katofsky 1993).

16) Membrane costs 68 $US\$_{1997}$/(kW/bar), but these costs are only 9 % of the total installed cost for a Hydrogen Separation Device. Investment costs stem from Parsons I&TG (1998). The economies of scale of the membrane surface are low because the required surface area is proportional to the throughput, this slightly influences the overall R factor of the complete HSD.

17) Costs for CO_2 removal through Selexol amounts 14.3 $MUS\$_{1993}$ fob (overall installation factor is 1.87) for an 810 kmol CO_2/hr unit, R=0.7 (Katofsky 1993) up to 44 $MUS\$_{1994}$ installed for a 9909 kmol CO_2/hour unit (Hendriks 1994). The value from Hendriks is assumed to be right, since his research into CO_2 removal is comprehensive.

18) FT Gas phase reactor operating at 40 bar costs 35 $MDfl_{2000}$ (direct costs + engineering) for a production of 100 MW_{FT} HHV (CO conversion = 150 mol/s, and $-R_{CO} = 6.03 \cdot 10^{-4}$ mol/s·kg_{cat}, thus reactor volume is 208 m^3 CNH), R = 1, there is no maximum scale, add 30 % to obtain total capital investment (Tijmensen et al. 2002). For operation at 60 bar and same volume, costs increase by 50 % (reactor with thin wall: thickness linear with pressure).

19) 19.995 $MUS\$_{1996}$ ISBL or 34.6 $MUS\$_{1996}$ "Total Costs" (includes offsites, HOService/Fee and contingency) here assumed to be total capital investment, for a FT slurry reactor train of *three* 362 m^3 reactors at 25.2 bar, each yielding 295 m^3_{FT}/day or 131 MW_{FT} (Bechtel 1996), the installation costs of one 362 m^3 reactor thus is 15.2 $M€_{2002}$. Operation at 60 bar makes the installation 2.4 times more expensive (see note 18).

20) Tijmensen (2002) assumes that a hydrocracking unit has installed costs (=Total capital investment, CNH) of 17.05 $MDfl_{2000}$ (8.22 $M€_{2002}$) per 2000 bbl/day or 11352 GJ_{C5+} per day (13.2 m^3/hr), but does not specify further, his source is confidential. According to Bechtel (1996) a wax hydrocracking plant of 475 Mlbs/hr of FT diesel+gasoline product (286 m_{FT}^3/hr) costs 71.5 $MUS\$_{1996}$ total capital investment, furthermore a distillate hydrotreater (29.1 $MUS\$_{1996}$), a naphtha hydrotreater (9.8), a naphtha reforming plant (49.3), a C_4 isomerisation plant (7.2), a C_5/C_6 isomerisation plant (11.2), a $C_3/C_4/C_5$ alkylation plant (41.2), and a saturated gas plant (9.2) are necessary, thus summing up to 233 $M€_{2002}$. Thus upgrading costs in Bechtel's study are 1.3 - 3.3 (at R = 1 - 0.7) times Tijmensen's values. Here we assume that the Bechtel numbers are right.

21) For indication: A complete Combined Cycle amounts about 830 $US\$_{1997}$ per installed kW_e. Quoted from (Solantausta et al. 1996) by (Oonk et al. 1997).

22) Scaled on Gas Turbine size.

23) Steam system consists of water and steam system, steam turbine, condenser and cooling. Scaled on Steam Turbine size.

24) Expansion turbine costs are assumed to be the same as steam turbine costs (without steam system).

25) Overall installation factor. Includes auxiliary equipment and installation labour, engineering and contingencies. Unless other values are given by literature, the overall installation factor is set 1.995 for a 70 MW_{th} scale (Faaij et al. 1998). This value is based on 33% added *direct costs* to installed equipment (instrumentation and control 5%, buildings 1.5% grid connections 5%, site preparation 0.5%, civil works 10%, electronics 7%, and piping 4%) and 50 % added *indirect costs* to direct costs (engineering 15%, building interest 10%, project contingency 10%, fees/overheads/profits 10%, start-up costs 5%). The added *direct costs* part of the overall installation factor decreases with scale (average R-factor is 0.82).

26) Maximum sizes from Tijmensen (2002), unless indicated otherwise.

Annex G Concept calculations data

Overview flows

Concept

Basic variations in gasification pressure and oxygen purity for wet routes, slurry and solid bed

Concept Summary

		1	2	3	4	5	6	7	8	9	10	11	12	13	14
Gasifier pressure	bar	1.3	1.3	1.3	6	6	25	25	1.3	1.3	1.3	6	6	25	25
Gasifier medium		20.75%	80.00%	99.50%	80.00%	99.50%	80.00%	99.50%	20.75%	80.00%	99.50%	80.00%	99.50%	80.00%	99.50%
Cleaning route		TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet
Reforming		0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%
Shifting		15%	2%	2%	2%	2%	2%	2%	15%	2%	2%	2%	2%	2%	2%
CO2 removal		0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%
FT reactor type		Slurry	Slurry	Slurry	Slurry	Slurry	Slurry	Slurry	Solid	Solid	Solid	Solid	Solid	Solid	Solid
FT reactor pressure	bar	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8
FT reactor dimensions	m3	3568.02	2496.10	2982.43	2690.11	3191.86	2883.55	3396.52	129.66	91.02	108.75	98.09	116.39	105.14	123.85
CO+H2 conversion		70%	70%	70%	70%	70%	70%	70%	70%	70%	70%	70%	70%	70%	70%
Alpha		0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.91	0.92	0.92	0.92	0.92	0.92	0.92
Degree of recycle		0	0	0	0	0	0	0	0	0	0	0	0	0	0
Short or long		no	no	no	no	no	no	no	no	no	no	no	no	no	no

Overall balance

	Biomass input	dry tonne/hour	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4
		MWth HHV	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0
	Fuel output	kg/hr light	1087.8	1204.8	1269.1	1222.9	1281.5	1239.7	1292.9	1282.8	1204.8	1268.9	1222.8	1281.3	1239.3	1292.9
		kJ/kg HHV	48051.0	48111.2	48120.0	48112.3	48120.3	48113.3	48120.5	48058.8	48111.2	48120.0	48112.2	48120.2	48113.2	48120.5
		kg/m3	614.8	622.3	623.4	622.4	623.4	622.5	623.4	615.4	622.3	623.4	622.4	623.4	622.5	623.4
		kg/hr heavy	8517.8	8344.1	8660.2	8452.5	8739.1	8551.2	8812.4	8151.7	8344.1	8660.2	8452.5	8739.1	8551.2	8812.4
		kJ/kg HHV	47508.0	47508.0	47508.0	47508.0	47508.0	47508.0	47508.0	47500.7	47508.0	47508.0	47508.0	47508.0	47508.0	47508.0
		kg/m3	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.9	6.7	6.7	6.7	6.7	6.7	6.7
		MW HHV	126.9	126.2	131.2	127.9	132.5	129.4	133.6	124.7	126.2	131.2	127.9	132.5	129.4	133.6
	Power Gas turbine	MWe	50.0	38.6	39.2	39.0	39.6	39.4	39.9	50.9	38.6	39.2	39.0	39.5	39.4	39.9
	Power Steam turbine	MWe	32.6	34.8	32.6	31.6	30.2	28.5	28.0	33.0	34.8	32.6	31.6	30.2	28.5	28.0
	Power needs Auxiliairies	MWe	37.5	28.2	26.8	17.4	16.6	12.2	11.3	37.5	28.2	26.8	17.4	16.6	12.2	11.3
	Power needs Oxygen production	MWe	0.0	6.5	6.7	5.9	6.1	5.4	5.6	0.0	6.5	6.7	5.9	6.1	5.4	5.6
	Net power output	MWe	45.1	38.7	38.2	47.3	47.1	50.2	51.0	46.4	38.7	38.2	47.3	47.1	50.2	51.0
	HHV efficiency		**43.0%**	**41.2%**	**42.4%**	**43.8%**	**44.9%**	**44.9%**	**46.2%**	**42.8%**	**41.2%**	**42.4%**	**43.8%**	**44.9%**	**44.9%**	**46.1%**

Stream and reactor dimensions

	Total Pre-treatment	wet tonne/hour	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21
Gasification system	CFB gasifier	dry tonne/hour	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43
	Air separation unit 99.5%	tonne O2/day			534.74		488.34		445.86			534.74		488.34		445.86
	Air separation unit 95%	tonne O2/day		520.69		475.42		433.94			520.69		475.42		433.94	
	Oxygen compressor	MWe				1.50	1.47	3.20	3.12				1.50	1.47	3.20	3.12
Gas Cleaning	Cyclones	m3 gas/s	153.06	147.27	130.47	29.64	26.32	6.63	5.91	153.06	147.27	130.47	29.64	26.32	6.63	5.91
	Tar cracker	m3 gas/s	156.18	150.27	133.13	30.24	26.85	6.77	6.03	156.18	150.27	133.13	30.24	26.85	6.77	6.03
	Oil scrubber	m3NTP gas/s														
	HT Heat Exchanger	MWth	107.66	116.24	104.08	111.98	100.83	108.76	98.40	107.66	116.24	104.08	111.98	100.83	108.76	98.40
	Particle filters	m3 gas/s	61.41	60.76	54.74	12.56	11.40	2.80	2.57	61.41	60.76	54.74	12.56	11.40	2.80	2.57
	Aqueous scrubbers	m3 gas/s	103.82	102.76	92.58	13.78	12.51	2.86	2.63	103.82	102.76	92.58	13.78	12.51	2.86	2.63
	Dry gas cleaning	m3 gas/s	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
	Guard beds (active C and ZnO)	m3NTP gas/s	49.65	29.56	28.44	29.43	28.37	29.31	28.32	49.72	29.56	28.44	29.43	28.37	29.31	28.32
Syngas Processing	Compressors	MWe	33.65	24.33	22.97	12.02	11.27	5.15	4.29	33.63	24.33	22.97	12.02	11.27	5.15	4.29
	Autothermal reformer	m3NTP gas/s														
	Shift Reactor	kmol CO+H2/hour	0.48	0.06	0.07	0.06	0.07	0.06	0.07	0.48	0.06	0.07	0.06	0.07	0.06	0.07
	PSA Units A + B	kmol total/hr	122.20	88.58	90.73	89.92	91.97	90.27	92.24	106.57	88.58	90.73	89.92	91.97	90.27	92.24
	Ceramic membrane	tonne H2/hr														
	Selexol CO2 removal	kmol CO2/hr														
FT Production	Solid bed gas phase FT	m3								129.66	91.02	108.75	98.09	116.39	105.14	123.85
	Slurry phase FT	m3	238.00	167.00	199.00	180.00	213.00	193.00	227.00							
	Hydrocracker	m3FT/hr	11.50	11.30	11.70	11.40	11.70	11.60	11.90	14.10	11.30	11.70	11.40	11.70	11.60	11.90
Power Generation	Gas turbine + HRSG	MWe	47.54	37.42	38.15	37.86	38.53	38.25	38.89	48.47	37.42	38.15	37.86	38.52	38.25	38.88
	Steam Turbine + steam system	MWe	32.58	34.83	32.57	31.64	30.24	28.52	28.00	33.01	34.83	32.58	31.64	30.23	28.52	27.98
	Expansion Turbine	MWe	2.47	1.17	1.03	1.14	1.02	1.12	1.01	2.47	1.17	1.03	1.14	1.02	1.12	1.01

Overview economics

			Concept													
			Basic variations in gasification pressure and oxygen purity for wet routes, slurry and solid bed													
			1	***2***	***3***	***4***	***5***	***6***	***7***	***8***	***9***	***10***	***11***	***12***	***13***	***14***
	Total Pre-treatment	M€	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0
Gasification system	CFB gasifier	M€	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6
	Air separation unit 99.5%	M€			34.3		32.0		29.9			34.3		32.0		29.9
	Air separation unit 95 %	M€		31.9		29.8		27.9			31.9		29.8		27.9	
	Oxygen compressor	M€				5.4	5.2	9.8	9.7				5.4	5.2	9.8	9.7
Gas Cleaning	Cyclones	M€	16.1	15.7	14.5	5.4	5.0	2.1	1.9	16.1	15.7	14.5	5.4	5.0	2.1	1.9
	Tar cracker	M€	27.2	26.2	23.3	6.6	6.1	2.5	2.3	27.2	26.2	23.3	6.6	6.1	2.5	2.3
	Oil scrubber	M€														
	HT Heat Exchanger	M€	13.0	13.5	12.7	13.3	12.5	13.0	12.3	13.0	13.5	12.7	13.3	12.5	13.0	12.3
	Particle filters	M€	10.2	10.1	9.5	3.9	3.7	1.6	1.5	10.2	10.1	9.5	3.9	3.7	1.6	1.5
	Aqueous scrubbers	M€	28.7	28.4	25.7	6.5	6.1	2.3	2.2	28.7	28.4	25.7	6.5	6.1	2.3	2.2
	Dry gas cleaning	M€	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
	Guard beds (active C and ZnO)	M€	0.4	0.3	0.3	0.3	0.3	0.3	0.3	0.4	0.3	0.3	0.3	0.3	0.3	0.3
Syngas Processing	Compressors	M€	47.8	36.6	34.9	20.6	19.5	10.3	8.9	47.8	36.6	34.9	20.6	19.5	10.3	8.9
	Autothermal reformer	M€														
	Shift Reactor	M€	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
	PSA Units A + B	M€	2.6	2.1	2.1	2.1	2.1	2.1	2.1	2.4	2.1	2.1	2.1	2.1	2.1	2.1
	Ceramic membrane	M€														
	Selexol CO2 removal	M€														
FT Production	Solid bed gas phase FT	M€								18.7	13.2	15.7	14.2	16.8	15.2	17.9
	Slurry phase FT	M€	24.7	19.1	21.7	20.2	22.8	21.2	23.8							
	Hydrocracker	M€	24.6	24.3	24.9	24.4	24.9	24.7	25.2	28.3	24.3	24.9	24.4	24.9	24.7	25.2
Power Generation	Gas turbine + HRSG	M€	64.8	55.3	56.0	55.7	56.4	56.1	56.7	65.6	55.3	56.0	55.7	56.4	56.1	56.7
	Steam Turbine + steam system	M€	25.1	26.3	25.1	24.7	23.9	23.0	22.7	25.3	26.3	25.1	24.7	23.9	23.0	22.7
	Expansion Turbine	M€	3.9	2.4	2.3	2.4	2.2	2.4	2.2	3.9	2.4	2.3	2.4	2.2	2.4	2.2
Total Investment		M€	412.6	415.8	410.9	344.8	346.4	322.9	325.5	411.3	409.9	404.9	338.8	340.5	316.9	319.5
Total Investment corrected for lifetime		M€	373.0	375.9	371.4	311.7	313.2	291.9	294.2	371.8	370.5	366.0	306.3	307.8	286.5	288.8
	Biomass input	tonne/hour	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4
	Biomass input	MW HHV	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0
	Load	hours	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0
	Biomass input	GJ/year	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5
Annual Costs	Capital	M€	48.9	49.2	48.7	40.8	41.0	38.2	38.5	48.7	48.5	47.9	40.1	40.3	37.5	37.8
	O&M	M€	18.2	18.3	18.1	15.2	15.2	14.2	14.3	18.1	18.0	17.8	14.9	15.0	13.9	14.1
	Biomass	M€	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6
	Costs/Income Power	M€	-10.8	-9.3	-9.2	-11.4	-11.3	-12.1	-12.2	-11.1	-9.3	-9.2	-11.4	-11.3	-12.1	-12.2
Total Annual Costs		M€	90.8	92.8	92.1	79.2	79.5	75.0	75.2	90.2	91.8	91.1	78.2	78.6	74.0	74.2
Production	Fuel output	MW HHV	126.9	126.2	131.2	127.9	132.5	129.4	133.6	124.7	126.2	131.2	127.9	132.5	129.4	133.6
	Power output	MWe	45.1	38.7	38.2	47.3	47.1	50.2	51.0	46.4	38.7	38.2	47.3	47.1	50.2	51.0
	Efficiency fuel	%	31.7	31.6	32.8	32.0	33.1	32.4	33.4	31.2	31.6	32.8	32.0	33.1	32.4	33.4
	Efficiency power	%	11.3	9.7	9.6	11.8	11.8	12.6	12.8	11.6	9.7	9.6	11.8	11.8	12.6	12.8
	Efficiency total HHV	%	43.0	41.2	42.4	43.8	44.9	44.9	46.2	42.8	41.2	42.4	43.8	44.9	44.9	46.1
Costs of fuel produced @ 400 MWth scale		**€/GJ HHV**	**24.83**	**25.53**	**24.37**	**21.51**	**20.85**	**20.11**	**19.54**	**25.12**	**25.27**	**24.11**	**21.24**	**20.60**	**19.85**	**19.29**

Finding the optimal conversion for 25 bar oxygen wet route																		*Variations to the key concept*			
15	***16***	***17***	***18***	***19***	***20***	***21***	***22***	***23***	***24***	***25***	***26***	***27***	***28***	***29***	***30***	***31***	***32***	***33***	***34***	***35***
25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25
99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%
TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+dry	OS+wet
0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	100%	0%	0%
2%	2%	2%	2%	15%	29%	36%	42%	45%	48%	55%	62%	69%	76%	39%	35%	32%	29%	25%	27%	9%
0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%
Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid
54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8
63.60	87.35	123.69	128.54	111.91	94.45	85.15	76.16	53.12	33.67	21.33	15.69	12.19	9.48	55.10	50.55	44.71	94.45	82.47	94.50	48.13
50%	60%	70%	71%	80%	90%	95%	99%	95%	90%	80%	70%	60%	50%	90%	80%	70%	90%	90%	90%	90%
0.92	0.92	0.92	0.92	0.92	0.90	0.88	0.87	0.86	0.85	0.83	0.79	0.75	0.70	0.88	0.89	0.89	0.90	0.89	0.90	0.87
0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
no	no	no	no	no	no	no	no	no	no	no	no	no	no	no	no	no	no	no	no	no
72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4
400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0
863.0	1074.3	1294.0	1316.3	1529.9	2537.3	3405.3	4271.1	4253.4	4152.7	3982.2	3650.3	3074.1	2182.7	3374.7	2580.4	2008.0	2537.3	2422.6	2547.6	2753.7
48079.5	48099.9	48120.7	48122.8	48146.0	48207.8	48260.8	48312.7	48287.7	48262.8	48237.8	48228.2	48224.9	48222.4	48234.4	48181.2	48147.5	48207.8	48218.0	48207.8	48822.2
687.4	685.2	682.9	682.7	680.2	673.9	668.7	663.8	666.3	668.8	671.5	672.7	673.3	673.9	671.4	676.8	680.3	673.9	672.9	674.0	6.8
6310.0	7564.0	8815.1	8940.1	10068.3	10183.4	9757.9	9199.0	8329.0	7397.5	5477.6	3739.2	2237.8	1052.3	8811.6	8339.9	7614.5	10183.4	8538.7	10150.8	3899.3
47497.9	47497.9	47497.9	47497.9	47497.9	47483.4	47472.8	47464.6	47460.8	47457.5	47451.8	47450.6	47456.4	47473.4	47468.5	47474.2	47478.6	47483.4	47478.8	47483.1	47465.3
743.2	743.2	743.2	743.2	743.2	743.4	743.6	743.9	743.9	744.0	744.0	743.7	742.8	741.4	743.8	743.6	743.5	743.4	743.5	743.4	743.8
94.8	114.2	133.6	135.5	153.3	168.3	174.3	178.6	166.9	153.2	125.6	98.2	70.7	43.1	161.4	144.5	127.3	168.3	145.1	168.0	88.8
60.6	50.3	39.9	38.8	28.9	18.8	12.6	5.2	15.1	23.8	38.7	52.5	66.1	79.7	21.0	31.7	41.5	18.8	17.1	18.8	75.1
36.4	32.2	28.0	27.5	19.2	10.8	0.0	0.0	0.0	0.0	10.6	14.0	17.3	20.5	0.0	14.0	18.9	10.8	23.6	13.6	18.1
11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	12.7	11.3	11.0
5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	5.6	7.6	5.6	6.4
80.1	65.6	51.0	49.5	31.2	12.7	-4.3	-11.6	-1.8	6.9	32.4	49.6	66.5	83.3	4.1	28.8	43.5	12.7	20.4	15.6	75.8
43.7%	**44.9%**	**46.2%**	**46.3%**	**46.1%**	**45.2%**	**42.5%**	**41.7%**	**41.3%**	**40.0%**	**39.5%**	**37.0%**	**34.3%**	**31.6%**	**41.4%**	**43.3%**	**42.7%**	**45.2%**	**41.4%**	**45.9%**	**41.1%**
85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21
72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43
445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	445.86	611.82	445.86	510.65
3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	3.12	4.28	3.12	3.57
5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	5.91	6.41
6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	6.03	
																				38.75
98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	98.30	101.88	76.64	34.45
2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57	0.00	0.00
2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	2.63	0.00	1.50
0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	4.85	0.00
28.48	28.41	28.33	28.32	32.56	37.32	39.74	41.81	42.86	43.91	46.32	48.70	51.06	53.39	40.80	39.46	38.47	37.32	33.84	53.72	25.60
4.29	4.29	4.30	4.30	4.30	4.30	4.29	4.29	4.28	4.28	4.28	4.27	4.27	4.27	4.29	4.29	4.29	4.30	4.56	4.28	3.57
																		41.61		
0.07	0.07	0.07	0.07	0.50	0.97	1.20	1.40	1.51	1.61	1.84	2.07	2.31	2.54	1.30	1.17	1.07	0.97	0.73		0.15
66.04	79.16	92.25	93.56	105.37	85.72	68.52	54.12	44.40	35.45	19.65	9.24	3.32	0.72	56.76	60.08	59.28	85.72	68.33		30.70
																			1.82	
63.60	87.35	123.69	128.54	111.91	94.45	85.15	76.16	53.12	33.67	21.33	15.69	12.19	9.48	55.10	50.55	44.71	94.45	82.47	94.50	48.13
8.49	10.18	11.86	12.03	13.55	13.70	13.12	12.37	11.20	9.94	7.36	5.03	3.01	1.42	11.85	11.22	10.24	13.70	11.48	13.65	5.24
59.21	49.10	38.90	37.86	27.96	18.05	12.09	5.06	14.53	22.94	37.44	50.97	64.34	77.64	20.21	30.65	40.28	18.05	16.39	18.12	74.12
36.35	32.18	27.97	27.53	19.23	10.81					10.64	13.99	17.26	20.50		14.00	18.89	10.81	23.64	13.60	18.13
1.36	1.19	1.01	0.99	0.90	0.71	0.47	0.17	0.54	0.83	1.22	1.51	1.78	2.03	0.77	1.05	1.24	0.71	0.70	0.72	0.96

Finding the optimal conversion for 25 bar oxygen wet route																	Variations to the key concept			
15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34	35
46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0
77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6
29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	29.9	37.9	29.9	33.1
9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	9.7	12.5	9.7	10.8
1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9	2.0
2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	2.3	
																				3.2
12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.3	12.6	10.7	6.9
1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	0.0	0.0
2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	0.0	1.5
0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	4.5	0.0
0.3	0.3	0.3	0.3	0.3	0.3	0.4	0.4	0.4	0.4	0.4	0.4	0.5	0.5	0.4	0.4	0.3	0.3	0.3	0.5	0.2
8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	9.4	8.9	7.7
																		44.7		
0.0	0.0	0.0	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.0		0.0
1.7	1.9	2.1	2.2	2.3	2.0	1.7	1.5	1.3	1.1	0.7	0.4	0.2	0.1	1.5	1.6	1.6	2.0	1.7		1.0
																			4.2	
9.2	12.6	17.9	18.6	16.2	13.7	12.3	11.0	7.7	4.9	3.1	2.3	1.8	1.4	8.0	7.3	6.5	13.7	11.9	13.7	7.0
19.9	22.6	25.1	25.4	27.6	27.8	26.9	25.9	24.1	22.2	18.0	13.8	9.6	5.7	25.1	24.1	22.7	27.8	24.5	27.7	14.2
74.9	66.2	56.8	55.7	45.7	34.3	26.4	15.0	29.8	40.1	55.3	67.8	79.1	89.5	36.9	48.5	58.1	34.3	32.2	34.4	86.8
27.0	24.9	22.7	22.5	17.7	12.1					12.0	14.4	16.5	18.5		14.4	17.5	12.1	20.3	14.1	17.1
2.7	2.5	2.2	2.2	2.1	1.8	1.4	0.7	1.5	2.0	2.5	2.9	3.2	3.5	1.9	2.3	2.5	1.8	1.8	1.8	2.2
327.9	323.3	319.4	319.1	304.3	284.4	261.5	246.9	257.1	263.0	284.5	294.4	303.2	311.5	266.2	291.0	301.6	284.4	341.4	287.9	317.3
296.4	292.2	288.8	288.5	275.1	257.1	236.4	223.2	232.4	237.8	257.2	266.1	274.1	281.6	240.6	263.1	272.6	257.1	308.6	260.3	286.8
72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4
400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0
8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0
11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5
38.8	38.3	37.8	37.8	36.0	33.7	31.0	29.2	30.5	31.2	33.7	34.9	35.9	36.9	31.5	34.5	35.7	33.7	40.4	34.1	37.6
14.4	14.2	14.1	14.0	13.4	12.5	11.5	10.9	11.3	11.6	12.5	13.0	13.3	13.7	11.7	12.8	13.3	12.5	15.0	12.7	14.0
34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6
-19.2	-15.7	-12.2	-11.9	-7.5	-3.0	1.0	2.8	0.4	-1.7	-7.8	-11.9	-16.0	-20.0	-1.0	-6.9	-10.5	-3.0	-4.9	-3.7	-18.2
68.6	71.3	74.2	74.5	76.5	77.7	78.1	77.4	76.8	75.6	73.0	70.5	67.8	65.2	76.8	74.9	73.1	77.7	85.1	77.6	67.9
94.8	114.2	133.6	135.5	153.3	168.3	174.3	178.6	166.9	153.2	125.6	98.2	70.7	43.1	161.4	144.5	127.3	168.3	145.1	168.0	88.8
80.1	65.6	51.0	49.5	31.2	12.7	-4.3	-11.6	-1.8	6.9	32.4	49.6	66.5	83.3	4.1	28.8	43.5	12.7	20.4	15.6	75.8
23.7	28.5	33.4	33.9	38.3	42.1	43.6	44.7	41.7	38.3	31.4	24.5	17.7	10.8	40.4	36.1	31.8	42.1	36.3	42.0	22.2
20.0	16.4	12.8	12.4	7.8	3.2	-1.1	-2.9	-0.4	1.7	8.1	12.4	16.6	20.8	1.0	7.2	10.9	3.2	5.1	3.9	19.0
43.7	44.9	46.2	46.3	46.1	45.2	42.5	41.7	41.3	40.0	39.5	37.0	34.3	31.6	41.4	43.3	42.7	45.2	41.4	45.9	41.1
25.13	**21.69**	**19.28**	**19.09**	**17.32**	**16.03**	**15.55**	**15.06**	**15.97**	**17.14**	**20.18**	**24.92**	**33.33**	**52.48**	**16.52**	**18.00**	**19.94**	**16.03**	**20.38**	**16.04**	**26.56**

						Recycles for 25 bar oxygen wet routes, both short and long						Influence of scale80, 400, 1000 and 2000 MWth input. For both wet and dry routes							
36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52	53	54	55
25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25	25
99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%	99.50%
TC+wet	OS+wet	OS+wet	OS+wet	TC+wet	TC+wet	TC+wet	TC+dry	OS+wet	TC+wet	TC+dry	OS+wet	TC+wet	TC+wet	TC+wet	TC+wet	TC+dry	TC+dry	TC+dry	TC+dry
0%	0%	100%	100%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%
28%	9%	35%	35%	29%	30%	0%	0%	0%	2%	2%	2%	29%	29%	29%	29%	27%	27%	27%	27%
98%	98%	0%	98%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%
Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid	Solid
54.8	54.8	54.8	54.8	54.8	22.5	23.0	23.0	23.0	22.5	23.7	22.2	54.8	54.8	54.8	54.8	54.8	54.8	54.8	54.8
95.19	45.93	100.77	98.77	92.58	96.89	92.23	85.71	30.98	275.21	282.39	103.86	94.45	18.93	236.28	470.42	94.50	18.90	235.58	469.21
90%	90%	90%	90%	90%	90%	70%	70%	70%	70%	70%	70%	90%	90%	90%	90%	90%	90%	90%	90%
0.92	0.90	0.90	0.92	0.90	0.75	0.76	0.75	0.67	0.80	0.81	0.76	0.90	0.90	0.90	0.90	0.90	0.90	0.90	0.90
0	0	0	0	0	0	0.5	0.5	0.5	0.5	0.5	0.5	0	0	0	0	0	0	0	0
no	no	no	no	no	no	short	short	short	long	long	long	no	no	no	no	no	no	no	no
72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	14.5	181.1	362.2	72.4	14.5	181.1	362.2
400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	80.0	1000.0	2000.0	400.0	80.0	1000.0	2000.0
1862.3	2652.4	2626.9	1995.4	2584.4	4788.4	4065.4	4016.8	2686.7	3950.8	3913.0	3266.0	2535.3	506.0	6334.2	12724.0	2547.6	509.6	6386.1	12823.1
48196.9	49017.6	48217.9	48217.6	48208.1	48234.8	48150.7	48150.0	48242.2	48131.6	48129.7	48244.2	48207.2	48207.1	48207.1	48207.3	48207.8	48207.8	48207.9	48208.1
675.0	4.8	672.9	672.9	673.9	672.4	680.9	681.0	672.4	682.6	682.8	671.8	674.0	674.0	674.0	674.0	674.0	673.9	673.9	673.9
11395.5	4727.7	10578.4	11639.9	10199.0	3249.8	3516.2	3298.8	834.2	4985.8	5166.4	1862.4	10183.3	2036.3	25456.9	50932.8	10150.8	2030.2	25383.9	50785.1
47497.9	47482.1	47485.1	47497.9	47482.8	47457.4	47456.2	47455.5	47483.0	47323.6	47316.3	47456.2	47483.4	47483.4	47483.4	47483.2	47483.1	47483.1	47483.0	47482.9
743.2	743.4	743.3	743.2	743.4	742.7	742.9	743.1	740.9	754.1	754.9	742.9	743.4	743.4	743.4	743.4	743.4	743.4	743.4	743.4
175.3	98.5	174.7	180.3	169.1	107.0	100.7	97.2	47.0	118.4	120.2	68.3	168.3	33.6	420.6	842.2	168.0	33.6	420.3	841.6
15.0	68.8	20.5	16.7	19.0	21.5	40.5	40.9	75.4	43.6	43.7	68.1	18.8	3.8	47.0	94.1	18.8	3.8	47.1	94.5
9.9	16.3	0.0	0.0	10.3	12.1	21.4	37.8	10.1	43.0	50.5	20.0	10.8	2.2	27.1	54.6	13.6	2.7	34.0	67.9
12.7	12.3	15.2	16.5	11.3	7.8	9.8	9.8	11.6	9.3	9.0	25.4	11.3	2.3	28.2	56.5	11.3	2.3	28.2	56.3
5.6	6.4	11.8	11.8	5.6	5.6	7.6	7.6	10.8	6.7	6.6	10.8	5.6	1.1	13.9	27.9	5.6	1.1	13.9	27.9
6.7	66.4	-6.6	-11.7	12.4	20.2	44.7	61.3	63.0	70.7	78.5	52.0	12.7	2.6	31.9	64.4	15.6	3.1	39.0	78.2
45.5%	**41.2%**	**42.0%**	**42.2%**	**45.4%**	**31.8%**	**36.3%**	**39.6%**	**27.5%**	**47.3%**	**49.7%**	**30.1%**	**45.3%**	**45.2%**	**45.2%**	**45.3%**	**45.9%**	**45.9%**	**45.9%**	**46.0%**
85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	85.21	17.04	213.03	426.06	85.21	17.04	213.03	426.06
72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	72.43	14.49	181.08	362.15	72.43	14.49	181.08	362.15
445.86	510.66	946.85	946.86	445.86	445.86	604.62	608.89	863.73	532.76	530.07	861.35	445.86	89.17	1114.66	2229.32	445.86	89.17	1114.66	2229.32
3.12	3.57	6.63	6.63	3.12	3.12	4.23	4.26	6.05	3.73	3.71	6.03	3.12	0.62	7.80	15.61	3.12	0.62	7.80	15.61
5.91	6.41	6.41	6.41	5.91	5.91	5.91	5.91	6.41	8.65	8.72	11.38	5.91	1.18	14.77	29.53	5.91	1.18	14.77	29.53
6.03				6.03	6.03	6.03	6.03		8.82	8.89		6.03	1.21	15.07	30.13	6.03	1.21	15.07	30.13
	38.75	38.75	38.75					38.75			68.79								
96.82	34.45	32.62	32.62	96.82	96.82	93.10	71.61	28.58	137.50	107.33	59.14	98.30	19.69	245.88	490.06	76.64	15.32	191.08	380.62
2.57	0.00	0.00	0.00	2.57	2.57	2.57	0.00	0.00	3.49	0.00	0.00	2.57	0.51	6.42	12.85	0.00	0.00	0.00	0.00
2.63	1.50	1.50	1.50	2.63	2.63	2.63	0.00	1.50	3.56	0.00	2.75	2.63	0.53	6.56	13.13	0.00	0.00	0.00	0.00
0.00	0.00	0.00	0.00	0.00	0.00	0.00	4.85	0.00	0.00	6.64	0.00	0.00	0.00	0.00	0.00	4.85	0.97	12.13	24.27
21.41	15.76	40.97	22.72	37.44	38.25	59.09	48.58	71.45	40.18	41.11	45.90	37.32	7.46	93.28	186.72	53.72	28.90	100.29	178.02
5.67	4.87	4.69	6.01	4.31	0.81	1.67	1.65	1.68	1.72	1.46	15.52	4.30	0.86	10.74	21.50	4.28	0.86	10.69	21.34
		25.53	25.53			14.97	15.38	18.23											
0.94	0.15	1.25	1.25	0.98	1.01	0.98		1.99	0.08		0.07	0.97	0.19	2.42	4.86				
118.69	50.26	97.01	128.71	84.63	4.52	9.42		0.76	292.65		4.79	85.72	17.17	214.39	427.54				
							1.92			0.17						1.82	0.36	4.55	9.12
1483.18	1423.09		1447.17																
95.19	45.93	100.77	98.77	92.58	96.89	92.23	85.71	30.98	275.21	282.39	103.86	94.45	18.93	236.28	470.42	94.50	18.90	235.58	469.21
15.33	6.36	14.23	15.66	13.72	4.38	4.73	4.44	1.13	6.61	6.84	2.51	13.70	2.74	34.24	68.51	13.65	2.73	34.15	68.32
15.02	68.79	20.48	16.67	18.97	20.48	39.46	39.75	74.07	42.63	42.81	66.94	18.08	3.61	45.18	90.57	18.12	3.62	45.35	90.89
9.91	16.33			10.35	12.11	21.45	37.84	10.06	43.04	50.50	20.04	10.82	2.17	27.09	54.58	13.60	2.72	33.99	67.91
0.20	0.37	0.73	0.22	0.71								0.71	0.14	1.78	3.56	0.72	0.14	1.80	3.59

						Recycles for 25 bar oxygen wet routes, both short and long						Influence of scale80, 400, 1000 and 2000 MWth input. For both wet and dry routes							
36	***37***	***38***	***39***	***40***	***41***	***42***	***43***	***44***	***45***	***46***	***47***	***48***	***49***	***50***	***51***	***52***	***53***	***54***	***55***
46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	46.0	13.8	105.3	205.9	46.0	13.8	105.3	205.9
77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	77.6	25.2	192.0	384.0	77.6	25.2	192.0	384.0
29.9	33.1	52.7	52.7	29.9	29.9	37.6	37.8	49.1	34.2	34.1	49.0	29.9	9.0	59.5	100.1	29.9	9.0	59.5	100.1
9.7	10.8	17.8	17.8	9.7	9.7	12.4	12.4	16.5	11.2	11.1	16.5	9.7	2.6	20.3	35.7	9.7	2.6	20.3	35.7
1.9	2.0	2.0	2.0	1.9	1.9	1.9	1.9	2.0	2.4	2.5	2.9	1.9	0.7	3.5	5.4	1.9	0.7	3.5	5.4
2.3				2.3	2.3	2.3	2.3		3.0	3.0		2.3	0.8	4.2	6.6	2.3	0.8	4.2	6.6
	3.2	3.2	3.2					3.2			4.8								
12.2	6.9	6.7	6.7	12.2	12.2	12.0	10.3	6.2	14.9	13.0	9.3	12.3	5.1	20.6	30.3	10.7	4.4	17.9	26.3
1.5	0.0	0.0	0.0	1.5	1.5	1.5	0.0	0.0	1.8	0.0	0.0	1.5	0.6	2.6	3.9	0.0	0.0	0.0	0.0
2.2	1.5	1.5	1.5	2.2	2.2	2.2	0.0	1.5	2.7	0.0	2.3	2.2	0.8	4.0	6.3	0.0	0.0	0.0	0.0
0.0	0.0	0.0	0.0	0.0	0.0	0.0	4.5	0.0	0.0	6.1	0.0	0.0	0.0	0.0	0.0	4.5	1.0	10.8	21.1
0.2	0.1	0.4	0.2	0.3	0.3	0.5	0.4	0.6	0.4	0.4	0.4	0.3	0.1	0.8	1.7	0.5	0.3	0.9	1.6
11.2	9.9	9.6	11.7	8.9	2.3	4.2	4.1	4.2	4.2	3.7	25.4	8.9	2.4	18.8	33.1	8.9	2.4	18.7	32.9
		34.5	34.5			26.1	26.4	28.9											
0.1	0.0	0.1	0.1	0.1	0.1	0.1		0.1	0.0		0.0	0.1	0.0	0.1	0.2				
2.5	1.4	2.2	2.7	2.0	0.3	0.4		0.1	4.8		0.3	2.0	0.7	3.8	6.2				
							4.4			0.6						4.2	1.2	8.8	15.3
16.7	16.2		16.4																
13.8	6.6	14.6	14.3	13.4	5.7	5.6	5.2	1.9	16.3	17.7	6.1	13.7	2.7	34.2	68.0	13.7	2.7	34.1	67.8
30.1	16.2	28.5	30.5	27.8	12.5	13.2	12.6	4.8	16.7	17.1	8.5	27.8	9.0	52.7	85.7	27.7	9.0	52.6	85.5
30.4	82.6	37.3	32.6	35.4	37.3	57.3	57.6	86.8	60.3	60.4	81.2	34.3	12.1	62.6	99.1	34.4	12.1	62.8	99.4
11.5	15.9			11.8	13.1	19.1	27.7	11.6	30.2	33.6	18.2	12.2	4.3	22.2	35.4	14.1	4.9	25.8	40.9
0.8	1.2	1.8	0.8	1.8								1.8	0.6	3.2	5.0	1.8	0.6	3.2	5.0
300.5	331.4	336.4	351.3	284.9	254.9	319.8	331.4	341.2	326.6	326.8	348.4	284.5	90.5	610.5	1112.7	287.9	90.8	620.4	1133.6
271.6	299.6	304.1	317.5	257.6	230.4	289.1	299.6	308.5	295.3	295.4	315.0	257.2	81.8	551.9	1005.9	260.3	82.1	560.8	1024.7
72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	72.4	14.5	181.1	362.2	72.4	14.5	181.1	362.2
400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	400.0	80.0	1000.0	2000.0	400.0	80.0	1000.0	2000.0
8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0	8000.0
11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	11.5	2.3	28.8	57.6	11.5	2.3	28.8	57.6
35.6	39.2	39.8	41.6	33.7	30.2	37.9	39.3	40.4	38.7	38.7	41.3	33.7	10.7	72.3	131.8	34.1	10.8	73.5	134.2
13.2	14.6	14.8	15.5	12.5	11.2	14.1	14.6	15.0	14.4	14.4	15.3	12.5	5.0	25.3	45.1	12.7	5.1	25.8	45.9
34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	34.6	6.9	86.4	172.8	34.6	6.9	86.4	172.8
-1.6	-15.9	1.6	2.8	-3.0	-4.9	-10.7	-14.7	-15.1	-17.0	-18.8	-12.5	-3.1	-0.6	-7.7	-15.4	-3.7	-0.7	-9.4	-18.8
81.8	72.5	90.8	94.4	77.9	71.1	75.8	73.7	74.9	70.6	68.8	78.7	77.7	22.1	176.4	334.2	77.6	22.0	176.3	334.2
175.3	98.5	174.7	180.3	169.1	107.0	100.7	97.2	47.0	118.4	120.2	68.3	168.3	33.6	420.6	842.2	168.0	33.6	420.3	841.6
6.7	66.4	-6.6	-11.7	12.4	20.2	44.7	61.3	63.0	70.7	78.5	52.0	12.7	2.6	31.9	64.4	15.6	3.1	39.0	78.2
43.8	24.6	43.7	45.1	42.3	26.7	25.2	24.3	11.8	29.6	30.1	17.1	42.1	42.0	42.1	42.1	42.0	42.0	42.0	42.1
1.7	16.6	-1.6	-2.9	3.1	5.1	11.2	15.3	15.8	17.7	19.6	13.0	3.2	3.2	3.2	3.2	3.9	3.9	3.9	3.9
45.5	41.2	42.0	42.2	45.4	31.8	36.3	39.6	27.5	47.3	49.7	30.1	45.3	45.2	45.2	45.3	45.9	45.9	45.9	46.0
16.20	**25.55**	**18.04**	**18.18**	**15.98**	**23.07**	**26.13**	**26.32**	**55.29**	**20.72**	**19.87**	**39.99**	**16.04**	**22.77**	**14.56**	**13.78**	**16.04**	**22.71**	**14.56**	**13.79**