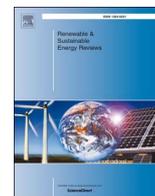




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Techno-economic and uncertainty analysis of Biomass to Liquid (BTL) systems for transport fuel production

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ABSTRACT

This work examines the technical and economic feasibility of Biomass-To-Liquid (BTL) processes for the manufacture of liquid hydrocarbon fuels. Six BTL systems are modelled and evaluated which are based on pressurised oxygen gasification of woody biomass, and specifically on circulating fluidised bed and entrained flow gasification systems. Three fuel synthesis technologies are considered: Fischer-Tropsch synthesis, methanol conversion followed by Methanol to Gasoline (MTG) and the Topsoe Integrated Gasoline (TIGAS) synthesis.

Published modelling studies of BTL systems based on gasification have only used deterministic estimates of fuel production costs to assess economic viability without accounting for uncertainties of their model parameters. Unlike other studies, the present techno-economic assessment examines and quantifies the effect of uncertainty of key parameters on the fuel production costs. The results of this analysis show that there is a realistic chance (8–14%) of concepts based on Fischer-Tropsch synthesis meeting the cost of conventional fuels; that this probability could be increased to 50% with moderate tax incentives (an 8% reduction in the tax rate); but that deterministic estimates may be systematically underestimating likely production costs.

The overall energy efficiency and production costs of the BTL designs evaluated range from 37.9% to 47.6% LHV and €17.88–25.41 per GJ of produced fuels, respectively. The BTL concept with the lowest production costs incorporates CFB gasification and FT synthesis. The model deterministic estimates of production costs of this design indicate that a BTL process is not yet competitive with conventional refineries since the biofuel production costs are approximately 8% higher than current market prices. Large scale biofuel production may be possible in the long term through subsidies, crude oil price rises and legislation.

1. Introduction

For the last four decades there has been a considerable interest in producing liquid transportation fuels from biomass as costs of petroleum continue to rise, which has been reinforced by subsequent environmental concerns. Since the Industrial Revolution, humans have significantly added to the amount of heat-trapping greenhouse gases in the atmosphere by burning fossil fuels that emit CO₂, cutting down forests that reduces CO₂ absorption and other activities (e.g. transporting goods and people, waste disposal). It is believed that the significant increase in anthropogenic greenhouse gas emissions since the beginning of industrial revolution (e.g. 40% increase for CO₂) is the main reason behind the observed rise in average global temperatures [1].

In addition to environmental concerns and according to the current facts, energy experts predict a 35% increase in worldwide petroleum

demand by 2025 [2]. This will increase dependency on a relatively limited number of oil producing countries with serious risks for energy security and global social stability. Regarding the oil market, it is predicted that the Middle East will continue to be in dominant position as it has the greatest proven oil reserves in the world. Conversely, nations with less petroleum resources will be vulnerable to energy shortages unless they develop alternative sources of energy. Such alternatives include nuclear, wind, solar, hydroelectricity, wave, tidal, geothermal and energy from biomass.

Biomass derived transport fuels (biofuels) can play an important role in reducing greenhouse gas emissions and dependency on fossil fuels by limiting or reducing consumption and combustion of fossil fuels [3]. This is also why the European Union has set ambitious targets for the application of biofuels through EU Biofuels Directive 2009/28/EC. According to the directive, 10% of all transport fossil fuels sold in EU countries, calculated on the basis of energy content, should be replaced

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with biofuels by 2020 [4].

Nowadays, the substitution of transport fossil fuels with biofuels is already feasible by state-of-the-art renewable liquid fuels, such as bioethanol for gasoline engines, produced by fermentation of sugar or starch and biodiesel for diesel engines produced via transesterification of vegetable oils or animal fats [5]. These so-called “first generation biofuels” are characterised by an unexpected growth following government subsidies and legislative pressures, however there are some serious problems associated with their application with respect to feedstock requirements and land availability – the food vs fuel debate. In addition to the consequences on economy and land competition, net carbon savings from first generation biofuels are questionable due to the clearance of virgin land (e.g. rain forests) for cultivation, high fertilizer requirement and low productivity per hectare [6].

In order to overcome the above mentioned shortages, the so-called “second generation biofuels” have been introduced. Unlike first generation biofuels, they avoid using food resources and also make use of a wider range of biomass feedstocks than just plant oils and sugar/starch components. These sources include non-food biomass, dedicated energy crops and biomass co-products and waste from many different sectors such as agriculture, horticulture, forestry, paper and pulp processing and wastes, such as MSW. [7].

This study examines the technical and economic feasibility of processes that manufacture second generation liquid fuels from non-food crops and wastes which are referred to as Biomass-To-Liquid (BTL) processes. The term “BTL” is only applied to thermo-chemical processes, such as pyrolysis and gasification, and thus it is not used for biochemical routes (e.g. fermentation) to biofuel production. The scope was limited to synthetic liquid hydrocarbons (diesel, gasoline and kerosene) as these can be readily incorporated and integrated with conventional markets and supply chains while alcohols (e.g. ethanol, methanol, mixed alcohols) and ethers (e.g. DME – dimethyl ether) have more limited short term prospects in the UK and European transport fuel infrastructures [8].

Large scale coal-to-liquid (CTL) and gas-to-liquid (GTL) processes have been commercialised for decades (e.g. Sasol and Shell plants). This is not the case with BTL processes with only a few plants built to date on pilot and demonstration scale: In the late nineties, Choren started operating a 1 MW_{th} BTL plant in Freiberg, Germany and planned to build a commercial plant with a capacity of 15,000 t/yr of fuel products before filing for insolvency in July 2011 [9]. NSE Biofuels Oy operated a 12 MW_{th} (656 t/yr of fuels) BTL demonstration plant in Finland from 2009 to 2011 which employed a circulating fluidised bed (CFB) gasifier developed by Foster Wheeler [10]. Plans were made to build a commercial plant with a projected output capacity of 100,000 t/yr but it was never constructed due to lack of public funding [11]. In 2010, five French partners and Uhde launched BioTfuel with two pilot plants currently on operation in France: a biomass pretreatment plant with a torrefaction unit in Venette and an entrained flow gasification and Fischer-Tropsch (FT) synthesis plant near Dinkirk [12]. It is currently planned to validate the techno-economic feasibility of the whole process chain by 2020 before moving on to industrial scale production. The Karlsruhe Institute of Technology (KIT) bioliq pilot plant with a capacity of 1 t/day has been in operation since 2014 and produces gasoline via DME using a process similar to the Topsoe Integrated Gasoline Synthesis (TIGAS) process. More information on the Choren and KIT Bioliq processes is provided in Section 2.4.

The environmental and socio-economic impacts of large scale BTL projects are not known with certainty as there is not an industrial plant currently on operation. BTL plants consume biomass as feedstock and thus it is expected to reduce GHG emissions with respect to fossil fuel processes, especially if forest waste is used [13]. Energy crops, like miscanthus, are typically grown close to the conversion plant to reduce transportation costs. This prompts the development of associated industries for biomass growing, collecting and transporting and thus large BTL facilities could significantly enhance the local economy [14].

The techno-economics of BTL processes is a heavily researched topic with the main aim being to support policy makers and businesses in their decision making by identifying the most economic process designs and the parameters (e.g. biomass price) that significantly affect the economic competitiveness of these technologies. Tijmensen et al. [15] evaluated the co-production of transport fuels and power from integrated biomass CFB gasification and FT synthesis. The cost of fuel products was estimated at 19.6 €₂₀₁₄ per GJ at a co-production efficiency of 45% (LHV) for oxygen blown pressurised gasification (2000 dry t/d plant capacity). Swanson et al. [16] modelled and compared two BTL process concepts based on entrained flow and CFB gasification. Both concepts included FT synthesis for the production of liquid fuels and electricity as a co-product. The entrained flow gasification concept resulted in higher biomass to fuel efficiencies and lower production costs compared to the CFB gasification design at 53% (LHV) and 27.1 €₂₀₁₄ per GJ, respectively. Boerrigter [17] also examined the economic competitiveness of entrained flow gasification for BTL production. The production cost was estimated at approximately 15.8 €₂₀₁₄ per GJ when the plant was scaled up to 9100 MW_{th}. Baliban et al. [13] evaluated BTL concepts based on other fuel synthesis options in addition to FT synthesis: the methanol-to-Gasoline (MTG) and the Mobil-Olefins-to-Gasoline/Distillate (MOGD) processes. The authors developed an optimization framework for the process synthesis of a BTL refinery and the economic feasibility of 24 BTL process designs was investigated. Production costs ranged from 11.56 to 24.55 €₂₀₁₄ per GJ for woody biomass (forest residues). All BTL concepts were claimed to be economically viable for crude oil prices above \$80 per bbl and for a biomass feedstock price below \$120 per dry tonne. Researchers from KIT [18–20] have carried out BTL techno-economic studies focusing on the KIT bioliq process. Production costs ranged from 25 (3.3 GW_{th} plant capacity) to 35 €₂₀₁₄ per GJ (1 GW_{th} plant capacity) which were higher than those reported by most studies discussed above. As a reference, the market price (without taxes) of conventional diesel and gasoline in 2014 was €16.2 and €16.6 per GJ, respectively [21].

In techno-economic feasibility studies of BTL plants, production costs are estimated using a number of technical and economic parameters which, among others, include product yields, capital costs and raw material costs. The values used for these parameters have a degree of uncertainty and thus are not known with absolute accuracy. This results in uncertainty in the model's output (i.e. production costs) and can be reduced through acquiring more data. However, even then, the modeller can never be entirely certain of their models' estimates particularly in the case of new plant projects and technologies, such as a BTL plant, as there is no experience of a real life plant. The above studies typically assess uncertainty using sensitivity analysis where the effect on biofuel production costs of changing key model parameters is determined.

While sensitivity analysis can show how variation in a single parameter can affect production cost, it does not take into account the effect of simultaneous variation of parameters. This lack can lead to a systematic bias in the estimation of costs. For example if two quantities can each independently vary by $\pm 50\%$, their product can be between 75% lower and 125% higher than an estimate based on the product mean values of the variables. As this range isn't symmetric, an estimate based on varying one parameter at a time would underestimate the likely value.

Even where a deterministic estimate of production cost is not systematically biased, it does not give us any information about the probability with which a particular cost level will be met. Baker & Shittu argue [28] that knowledge of the probability distributions underlying estimates are “particularly important for determining near term optimal technology policy” and that, in the context of climate change damage, such knowledge can have a major impact on climate change technology policy, in some cases justifying significantly higher levels of R&D investment [29]. Similarly, Mills et al. argue [30] that investors are unwilling to make energy-related investments because of a

lack of quantification of risk and volatility, which leaves them unable to properly assess the risk-reward trade-off. As a lack of investment and policy support were cited as contributing factors to the insolvency of Choren and the failure of NSE Biofuels to build a commercial plant, it is clear that this problem needs to be addressed.

To control for the effect of simultaneous parameter variation and to allow for the quantification of risk, this paper includes an uncertainty analysis of biofuel production costs based on the Monte Carlo statistical method which is the most widely used means for uncertainty analysis studies [31,32]. Uncertainty analysis employs probabilistic descriptions of model inputs which propagate through the model and therefore, unlike sensitivity analysis, focuses on risk quantification, allowing us to derive probability distributions of model outputs

The technologies involved in converting biomass into liquid hydrocarbon fuels, as well as the respective process designs, are presented in the next section. Sections 3 and 4 outline the methodologies for the process modelling and economic assessment, respectively. The results of the study, including the results of the sensitivity and uncertainty analyses, are presented and discussed in Section 5 followed by the conclusions in Section 6.

2. Process description

Biomass-to-Liquid (BTL) is a multi-step process that converts biomass to liquid biofuels through thermochemical routes, such as pyrolysis and gasification. This study examines six different BTL plant concepts which consist of several discrete steps:

1. Reception, storage, handling (RSH) and preparation
2. Biomass gasification
3. Gas cleaning and conditioning to derive the correct gas quality and composition
4. Fuel synthesis with either FT synthesis, MTG or TIGAS

Pyrolysis routes to transport fuels were not considered (i.e. gasification of pyrolysis products, bio-oil upgrading technologies) due to unfavourable economics [18,33] and the early stages of development of these technologies compared to gasification routes. This section discusses the selection and combination of the technologies involved in the main conversion steps to construct the BTL concepts considered and provides the definition of the selected plant scale and biomass feedstock.

2.1. Reception, storage, handling (RSH) and preparation

Wood is the feedstock of choice in proposed biofuel systems because of its homogeneity, consistency and quality. Wood has also been and currently is used in commercial bioenergy plants [9,34–36]. There is more information available about wood production, handling and processing than any other biomass feedstock. Feed handling and preparation of wood biomass is a well-known process as it is already widely practised in the pulp and paper industry. For these reasons this study focuses exclusively on wood biomass and specifically wood chips since the majority of other techno-economic studies of BTL plants use wood chips as feedstock [15,17,22,25–27]. A daily throughput of 2016 t of dry biomass is set as the desired plant size for all process concepts, which is consistent with plant sizes considered in previously published studies and thus permits comparisons [15,16,22,25,26].

Biomass can be transported by road, rail or water. Road transport (trucks) is the typical mode of transport in bioenergy facilities since it offers flexibility and is particularly suited to facilities where the material is transported over distances of less than 100 km [37]. For BTL plants, low cost feedstock is preferred as it results in lower production costs. This usually limits transport distances to less than 100 km [38]. Therefore, it is assumed that wood chips are delivered to the plant via trucks [38]. The trucks are weighed as they enter the plant and the

wood chips are dumped into a storage pile. The wood chips are reclaimed from storage and conveyed to a magnetic separator and then screened to keep particle sizes within appropriate limits and prevent contamination of the feedstock by metal or rocks. The characteristics and costs of wood chips RSH have been thoroughly investigated by Toft [38] and Rogers [39].

Biomass drying is either by hot air (rotary dryer) or steam (superheated steam dryer). Air rotary dryers are the most common technology for biomass drying since they are less sensitive to particle size and have bigger capacity compared to other dryers [40]. However, rotary dryers are associated with higher fire risks since they have the longest retention times [40]. Superheated steam dryers (SSD) are less common but are safer than rotary dryers with respect to fire hazards. Since oxygen is not present in superheated steam dryers, the fuel cannot burn, even at elevated temperatures [40,41]. Fuel synthesis processes, such as FT synthesis and MTG generate significant amounts of steam which can be used for other processes within the plant. In this case, superheated steam drying could be preferable.

A grinder (hammer mill) is placed after the dryer to reduce the wood chips size to 1 mm [42,43] for the entrained flow gasification concepts. Contrary to entrained flow gasifiers, circulating fluidised bed gasifiers are capable of handling a wider variety of biomass particle sizes [3]. Thus no grinding would be required for the concepts based on circulating fluidised bed gasification. During drying, light biomass particles can get entrained in the gas flow of the dryer when significant biomass losses can occur. Consequently, the grinder is placed after the dryer in the feed preparation chain which also reduces power consumption and improves product consistency.

2.2. Biomass gasification

The two gasification technologies best suited for large-scale BTL plants are the circulating fluidised bed (CFB) and entrained flow (EF) gasification [3,8,17,44]. For circulating fluidised bed gasifiers, operating temperature varies between 700 and 1100 °C. CFBs require a large minimum size for viability of typically above 15 t/h dry feed rates and they are relatively easy to scale up from a few MWth to ~100 MWth [3]. Even for capacities above 100 MWth, there is confidence that the industry would be able to provide reliable gasifiers [3]. There is also considerable experience with CFB gasification of biomass (e.g. Varnamo plant, HTW, IGT). Entrained flow reactors are only potentially viable above around 20 dry t/h feed rate and have good scale-up potential [3]. In addition to its scalability, EF gasification has the advantage that extensive experience is available from coal entrained flow gasification plants (e.g. 2000 t/d coal-fired Shell gasifier in Buggenum, the Netherlands) that have been developed to substantial commercial scale units [45,46]. EF gasifiers can operate at much higher gasification temperatures of up to 1200–1400 °C. These high temperatures result in higher carbon conversion, very low tar and methane content and thus lower gas cleaning requirements compared to other gasifiers [16,17,42]. Therefore, these two gasification technologies were selected for evaluation in this study.

Both gasifiers are oxygen-blown and pressurised. Many authors [15–17,23,42,47] recommend oxygen-blown pressurised gasification for the production of synthetic hydrocarbon fuels from biomass. Oxygen-blown gasification is preferred to air-blown gasification since in the latter, nitrogen acts as an inert diluent, decreasing the efficiency of other processes and necessitating larger and thus more costly equipment. Conversely, oxygen gasification is associated with increased capital costs and energy requirements due to the need for air separation. However, the use of oxygen is usually justified by increased liquid fuels yield due to the higher syngas quality [15–17,23,42,47]. Capital costs are also decreased due to smaller equipment sizes resulting from the absence of nitrogen.

Pressurised gasification holds the advantage of avoiding a costly compression step before the synthesis process required in atmospheric

pressure systems and thus reducing complexity and costs. However, capital and operating costs of such plants can be increased due to the additional costs of building pressure vessels [3]. These additional costs are to some extent balanced by savings from reduced piping sizes and the avoidance of gas compressors for the synthesis reactor and higher efficiencies [3]. The circulating fluid bed demonstration plant at Värnamo in Sweden, which was built and operated by Foster Wheeler and Sydkraft is an example of pressurised gasification [34].

Biomass feeding for pressurised gasifiers is a technology area where additional development is needed. Lock hoppers are the conventional pressurising technology for coal-fired gasifiers. The main drawback of this pressurised feed system is the high requirements of inert gas (usually N_2 or recovered carbon dioxide) and the additional costs of the inert gas compression [42,48]. A number of efforts have been made to develop alternative pressurised feed systems to address the disadvantages of lock hoppers. These include rotary, screw, piston and screw/piston feeders [48]. None of these have yet been demonstrated in large-scale operations [48].

At the demonstration plant at Värnamo, the preconditioned feedstock was pressurised in a lock hopper system by N_2 before being passed to a CFB gasifier via a pressurised vessel [49]. While this is acceptable in a small-scale demonstration facility, such as the Värnamo plant, in a large-scale commercial plant this will be far too expensive and needs to be solved in another way [50]. Therefore, it was decided to use CO_2 as the pressurisation gas which is acquired from the downstream CO_2 removal unit. In this way nitrogen dilution in downstream equipment is also avoided. The use of CO_2 as pressurisation medium is also proposed by Swanson [16], Larson [23] and van Drift [42].

2.3. Gas cleaning and conditioning

Gas cleaning is considered to be one of the biggest challenges to the development of a successful BTL plant. The impurities in syngas need to be reduced to the level demanded by the catalytic fuel synthesis processes. For each plant, and also for each type of catalyst, the acceptable levels of contaminants may be different. Even though it is claimed that there are no insuperable problems associated with cleaning of biomass-based syngas, there is no large scale or long term experience.

Tars are the major gas cleaning issue in biomass gasification. Tars are condensable organic compounds of high molecular weight with boiling points ranging from 80 to 350 °C [47]. When the temperature in the system decreases to below 350 °C, tars start to condense in exit pipes and on filters resulting in blockages and clogged filters and ultimately in system failure [51]. For the CFB concepts, a catalytic autothermal steam reformer is employed for tar destruction, as well as reforming of light hydrocarbons and ammonia [52]. A tar cracker is not necessary for the EF concepts due to the high operating temperature of the gasifier (1400 °C) that inhibits tar formation.

Catalytic tar cracking is recommended by many authors [15,22,26,27,53], even though this technology has not yet been demonstrated on large-scale. However, there is a significant ongoing research on tar reforming catalysts, especially on dolomite and nickel-based catalysts. Specifically, at the Värnamo plant, 95–99% of tars in gas streams were successfully cracked in a catalytic (dolomite catalyst) cracker at 750–900 °C under laboratory conditions [54]. Other catalysts also have been proved effective at tar decomposition. Pfeifer achieved an almost complete tar destruction (approximately 98%) and considerable ammonia decomposition (approximately 40%) using commercial steam reforming nickel catalysts at temperatures above 850 °C [55]. These catalysts were tested in a laboratory scale reactor fed by slip streams taken from the dual fluidised bed steam gasifier plant in Güssing, Austria.

Filters and cyclones are employed for the removal of particulates and alkali compounds. For the EF concepts, a water quench is placed after the gasifier for removal of particulates and cooling of syngas. This is also proposed by several authors [25,26,43,56]. For the CFB gasifier,

cooling of syngas is done by a heat exchanger to recover the steam generated for tar cracking/reforming. A water quench is also used to remove ammonia from the syngas but for gasifiers coupled with a tar reformer, such as the CFB gasifier in this study, most of nitrogen compounds can be cracked in the tar reformer, thus a water quench is not employed for the CFB concepts.

Sulphur in the biomass mostly forms hydrogen sulphide (H_2S) with small amounts of carbonyl sulphide (COS). A Rectisol unit is considered for the removal of sulphur species, as well as CO_2 . Rectisol is a very efficient process for the removal of the acid compounds since the achieved concentration of CO_2 and H_2S can be as low as 2 ppm and 0.1 ppm respectively [45]. In addition, it has been proven successful in large scale coal gasification and FT synthesis plants (e.g. Sasol facilities in South Africa) [57]. Rectisol was also chosen due to data availability, as compared to other large scale acid gas removal processes, such as Amisol and Sulfinol. The low nitrogen, chloride and sulphur concentrations in the raw biomass syngas from oxygen-blown gasifiers ensure that the resulting HCN, NO_x and SO_2 emissions are well below permitted levels for the downstream fuel synthesis process. Nevertheless, Rectisol can also remove trace components, such as HCN and HCl [45].

Initial simulations of the EF concept showed that the H_2/CO molar ratio of the dust free syngas was lower than the required ratio ($H_2/CO = 2$) for FT and methanol synthesis. Therefore, the ratio was adjusted by using a water-gas-shift (WGS) reactor. Locating the WGS unit here allowed carbon dioxide produced in the shift reaction to be removed soon after in the Rectisol unit. Initial simulations of the CFB concept also showed that a WGS unit was not necessary for the CFB-based concepts since the H_2/CO ratio of the syngas exiting the tar cracker was approximately 2. This is mainly due to the fact that light hydrocarbons and tars are almost fully reformed in the tar cracker/reformer. This also depends on the composition of the raw syngas from the gasifier. EF gasifiers produce syngas with high concentrations of hydrogen and carbon monoxide due to the high conversion of tars and light hydrocarbons. This results in lower H_2/CO ratios compared to fluidised bed gasifiers.

2.4. Fuel synthesis

Liquid fuels were produced from syngas using either FT synthesis, methanol synthesis followed by the MTG process, or the TIGAS process. These three processes are the best developed syngas conversion technologies for transport fuel production. FT synthesis has already been used in large-scale coal-to-liquid (CTL) and gas-to-liquid (GTL) plants throughout the world [58,59]. Both the MTG and the TIGAS technologies have been successfully proven at demonstration scale plants [60,61].

2.4.1. Fischer-Tropsch synthesis

Fischer-Tropsch synthesis is a process for catalytically converting a mixture of carbon monoxide and hydrogen to a variety of organic compounds, mainly hydrocarbon products of different chain lengths, typically from C1 to C100. Among the most widely known fuel synthesis plants in the world are the CTL Fischer-Tropsch plants operated by Sasol in South Africa. The Sasol plant in Secunda is the world's largest CTL production facility producing 160,000 bbl/d of liquid hydrocarbon fuels [58]. This represents approximately 27% of South Africa's total liquid fuel production [58]. The Pearl GTL is the largest GTL implementation of FT synthesis and is located in Qatar. This plant is owned by Shell and produces 140,000 bbl/day of hydrocarbon fuels [59].

Choren was one of the leading European companies in the BTL technology based on Fischer-Tropsch synthesis. CHOREN's pilot facility, the Alpha Plant, was constructed in 1997 and had seen 17,000 operating hours by the end of 2004 [9]. Initially constructed for gasification trials, the 1 MW_{th} Carbo-V gasifier, which used forestry wood as

Table 1
The BTL process concepts analysed in the study.

BTL concept	Preparation	Gasification	Fuel synthesis	Fuel product
EF-FT	SSD dryer, grinder	Entrained flow	Fischer-Tropsch	Diesel, gasoline, kerosene
EF-MTG	SSD dryer, grinder	Entrained flow	MTG	Gasoline
EF-TIG	SSD dryer, grinder	Entrained flow	TIGAS	Gasoline
CFB-FT	SSD dryer	CFB	Fischer-Tropsch	Diesel, gasoline, kerosene
CFB-MTG	SSD dryer	CFB	MTG	Gasoline
CFB-TIG	SSD dryer	CFB	TIGAS	Gasoline

feedstock, was complemented with a Fischer-Tropsch reactor in 2002 and thereafter began to produce diesel from for automotive and research purposes [9,62]. Previously, the world's first commercial BTL plant was under construction by Choren in Frieberg Saxony. It was initially scheduled to be completed by late 2009 and it would have an output capacity of 15,000 t per year requiring an estimated 67,500 t per year of dry biomass (forestry wood and wood residues) [9,63]. However, Choren Industries filed for insolvency in July 2011 because of financial difficulties in starting up the new plant in Frieberg [64].

2.4.2. Methanol-to-Gasoline (MTG) synthesis

In the MTG process, methanol is first synthesised using well established commercial technology then converted to hydrocarbons and water over zeolite catalysts (ZSM-5) [65]. The MTG process, developed by Mobil (today: ExxonMobil) scientists in the 1970s, was the first major synthetic fuels development in the fifty years since the development of the FT process [60]. A Mobil MTG plant was operated in Motunui, New Zealand from 1985 to 1997 and produced 14,500 bbl/d of unleaded gasoline composed mainly of isoparaffins and aromatics with low benzene content and essentially zero sulphur [66]. The Motunui plant was designed to meet one-third of New Zealand's demand for transport fuels, thus lessening their dependence on fossil fuels import [67]. In 1997, the production of gasoline was abandoned due to poor economics but the plant continued to manufacture methanol [66].

Since the first MTG plant in New Zealand, ExxonMobil made improvements in the MTG technology in the early 1990s that reduced both capital investment (by 15–20%) and operating costs mainly due to reductions in the size and number of heat exchangers [68,69]. The first coal-to-gasoline MTG plant, utilising this second generation MTG technology, was constructed by Jincheng Anthracite Mining Group (JAMG) in China [68,69]. The plant started up in June 2009 and its current capacity is 2500 bpd [70].

2.4.3. Topsoe Integrated Gasoline Synthesis (TIGAS)

The main principle of the TIGAS technology is the incorporation of the methanol synthesis and the DME synthesis into a combined methanol and DME synthesis process, thus intermediate methanol production and storage are eliminated. It was developed by Haldor Topsoe to reduce investment costs and subsequently production costs of gasoline produced from the MTG process [61]. It was demonstrated in Houston Texas where natural gas was used as feed to the process. The demonstration plant of 1 t per day gasoline started up in early 1984 and terminated in January 1987 after 10,000 h of operation [61].

The bioliq process developed by Karlsruhe Institute of Technology (KIT) allows conversion of lignocellulosic biomass (e.g. residual straw) to gasoline via DME using a process similar to TIGAS. The bioliq pilot plant with a capacity of 1 t/day of gasoline has been in operation since 2014 and incorporates the following processing steps: decentralised fast pyrolysis to produce a pyrolysis bio-oil/char slurry, high-pressure entrained flow gasification of the pyrolysis slurry, hot gas cleaning, DME synthesis and gasoline synthesis [71].

Even though there is very limited available data on the TIGAS process, it was decided to evaluate this technology to compare it with the MTG process as the TIGAS process can result in lower capital costs compared to the MTG process. This was achieved by avoiding a costly

syngas compression prior to methanol synthesis, as well as due to the integration of the methanol synthesis and the DME synthesis into a combined oxygenate synthesis process.

3. Process modelling

The modelling task of this techno-economic study is the development of a steady-state representation of the several BTL process concepts discussed in the previous section. The purpose is to calculate mass and energy balances and thus overall efficiencies for each process design to enable capital and production cost estimates and thus comparisons of the selected BTL concepts.

3.1. Selected process concepts

Table 1 summarises the BTL process concepts selected for techno-economic evaluation in this work and provides their name abbreviations used throughout the paper. All process designs were modelled using the equation oriented process simulation software IPSEpro in order to determine mass balances, energy balances, and product distributions.

3.2. Process simulation software

IPSEpro is an equation oriented process simulation software which is licenced by SimTech Simulation Technology, an Austrian company located in Graz [72]. The standard IPSEpro package provided by SimTech already contains a model library for modelling conventional power plant processes (Advanced Power Plant Library). However, this library does not contain any models of gasifiers, driers, gas cleaning equipment, and synthesis reactors. For the simulation of biomass gasification and related processes, a special model library called Pyrolysis and Gasification Process Library has been developed by several researchers at the Vienna University of Technology [73,74] and is licenced by SimTech. The structure of this library has been expanded by Pröll [74] in order to include biomass-related substances (e.g. tars) and to cover inorganic solids (e.g. CaO, K₂O). It also includes models of gasifiers, steam reformers, cyclones, filters, etc. The models contain mass and energy balances and specific equations describing chemical conversions, splitting conditions, empiric correlations from measurements of real gasification plants, etc.

3.3. Biomass feedstock

For all process concepts throughout this study the same type of biomass is used for consistency. The biomass model is based on wood chips, as discussed previously. The elemental composition of the wood chips feedstock is shown in Table 2.

In this paper, plant capacities are expressed in dry tonne per day (dry t/d), where “dry” means 0% water content in the biomass feedstock and 1 t is 1000 kg. Unless otherwise stated mass yields and energy efficiencies are quoted on a dry ash-free (daf) biomass feed basis where the presence of water and ash in the feedstock are not taken into account for the purposes of the calculation.

Table 2
Wood chips characteristics [18].

Moisture content	30%
Ash content	1%
Elemental analysis (dry)	
C	52%
H	6.3%
O	40.32%
N	0.3%
S	0.05%
Cl	0.03%
LHV (dry)	19.7 MJ/kg
LHV (wet)	13.1 MJ/kg

3.4. Gasification

Fig. 1 shows the CFB and EF gasification concepts with their downstream gas clean-up configurations. The wood chips are dried from 30% to 10% [54] in a SSD dryer using superheated steam (200 °C, 12 bar) [75]. The dried biomass is then pressurised in a lock hopper system and fed to the pressurised oxygen-blown gasifiers. As discussed above, CO₂ produced from the Rectisol unit was used instead of N₂ as inert gas for the lock hopper system. Higman et al. [76] report inert gas requirements of 0.09 kg/kg dry biomass for pressurised gasifiers (~25 bar). This results in a 180 t/d CO₂ requirement for the lock hopper system of both gasifiers.

Oxygen at 95% purity and steam are fed into the gasifiers operating at a pressure of 28 bar and temperatures of 870 °C for the CFB and 1400 °C for the EF gasifier, respectively [16]. For both CFB and EF

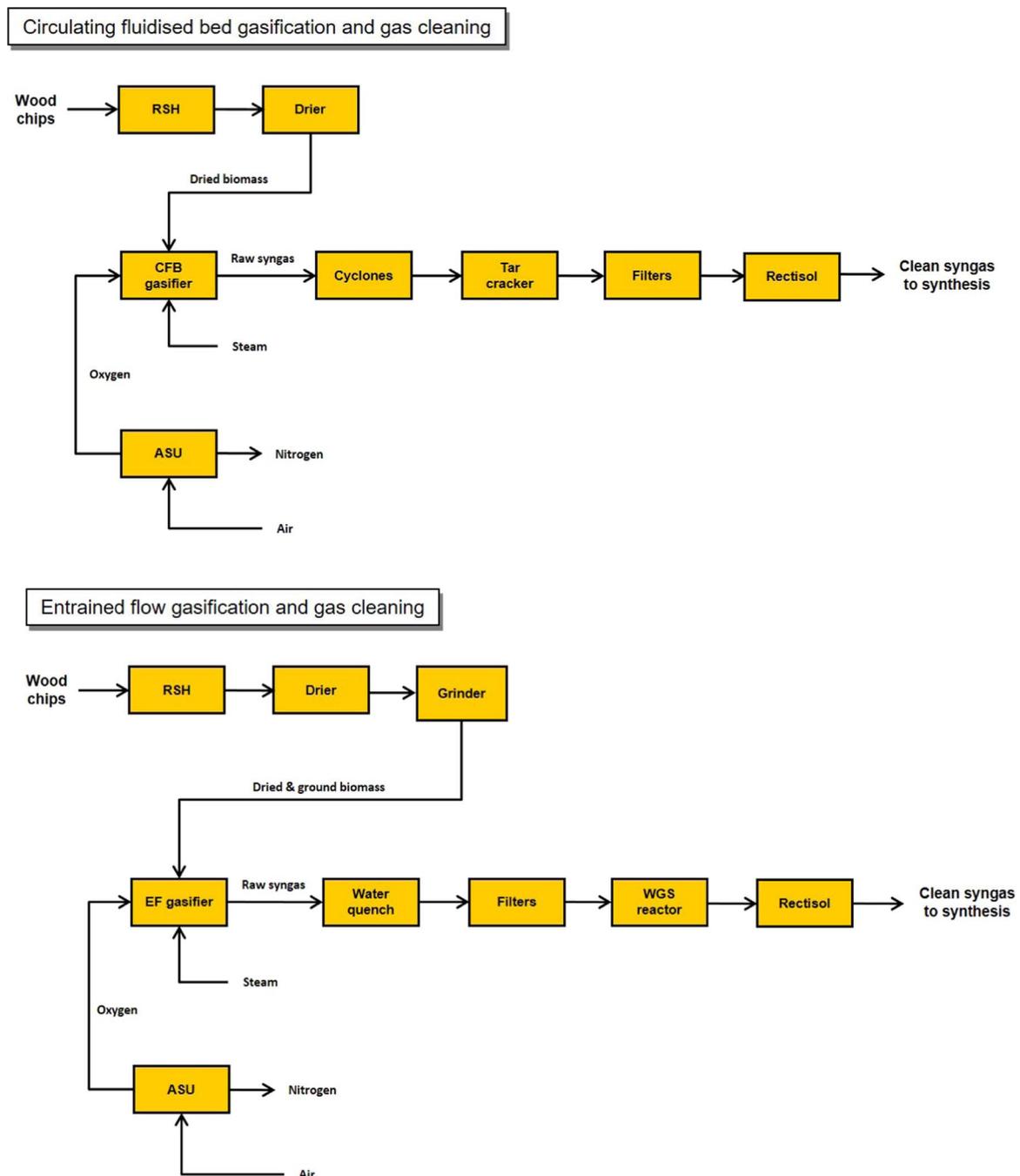


Fig. 1. Block flow diagrams of the circulating fluidised bed (CFB) and entrained flow (EF) gasification concepts. RSH: reception, storage, handling. WGS: water-gas-shift.

Table 3
Gasification process characteristics and raw syngas composition (vol%).

	CFB gasifier	EF gasifier
P (bar)	28	28
T (°C)	870	1400
Oxygen (kg/kg dry feed)	0.32	0.6
Steam (kg/kg dry feed)	0.17	0.15
Gas composition (vol% wet basis [dry basis])		
H ₂ O	12.6 [0]	25 [0]
H ₂	28.3 [32.4]	25.9 [34.5]
CO	26 [29.8]	37.1 [49.5]
CO ₂	21.2 [24.2]	10.8 [14.4]
CH ₄	10.5 [12]	0 [0]
C ₂₊	0.52 [0.6]	0 [0]
Ar	0.27 [0.3]	0.42 [0.55]
N ₂	0.56 [0.62]	0.75 [0.99]
NH ₃	0.005 [5.8 × 10 ⁻³]	0 [0]
H ₂ S	0.02 [0.024]	0.017 [0.023]
HCl	0.01 [0.013]	0.009 [0.013]
HCN	5 × 10 ⁻⁴ [6 × 10 ⁻⁴]	0 [0]

gasifiers, the amount of oxygen entering the gasifier was not set, as it was calculated by IPSEpro from the operating temperature of the gasifier. The amount of steam was adjusted accordingly in order to have a syngas composition similar to reported experimental results [45,77,78]. Table 3 shows the resulting syngas composition for both gasifiers calculated by IPSEpro. More hydrogen and carbon monoxide are formed from the entrained flow gasifier as a result of the water-gas-shift reaction and the reforming of light hydrocarbons. In general, according to

Le Chatelier's principle, higher temperatures favour the reactants in exothermic reactions (e.g. Water-gas-shift reaction) and favour the products in endothermic reactions (e.g. steam reforming reaction). The CFB gasifier, on the other hand, produces tar and a significant amount of methane and other light hydrocarbons, thus requiring downstream reforming.

3.5. Gas cleaning and conditioning

For the CFB gasification concepts, after the initial particulates separation by a cyclone, the syngas passes to the tar cracker where tars are destroyed at 875 °C by addition of oxygen and steam. It is assumed that light hydrocarbons (C₁-C₃) are converted at 99% conversion to syngas [54,55,79,80]. Tars are assumed to be fully converted into gaseous compounds [23] and all gases are determined in the model via elemental mass balances. The steam to carbon ratio of the tar cracker model was set at 2 for higher conversion of light hydrocarbons as proposed by Zeman and Hofbauer [81]. The tar free syngas is then cooled to 280 °C by a heat exchanger yielding steam which is used by the tar cracker. The cooled syngas passes through a bag filter [45] where the remaining particulates are removed. After the final particulate removal, the syngas is fed to the Rectisol unit where CO₂ and sulphur compounds are removed. In order to avoid catalyst poisoning, sulphur must be removed to at least 1 ppm by volume before the gas passes to the fuel synthesis process [45,47,82]. Rectisol can efficiently remove acid compounds in the syngas as the achieved concentration of CO₂ and H₂S could be as low as 2 ppm and 0.1 ppm by volume, respectively [45]. The Rectisol model was designed to leave a CO₂ volume fraction of 2% and a H₂S concentration of 0.1 ppm by volume in the

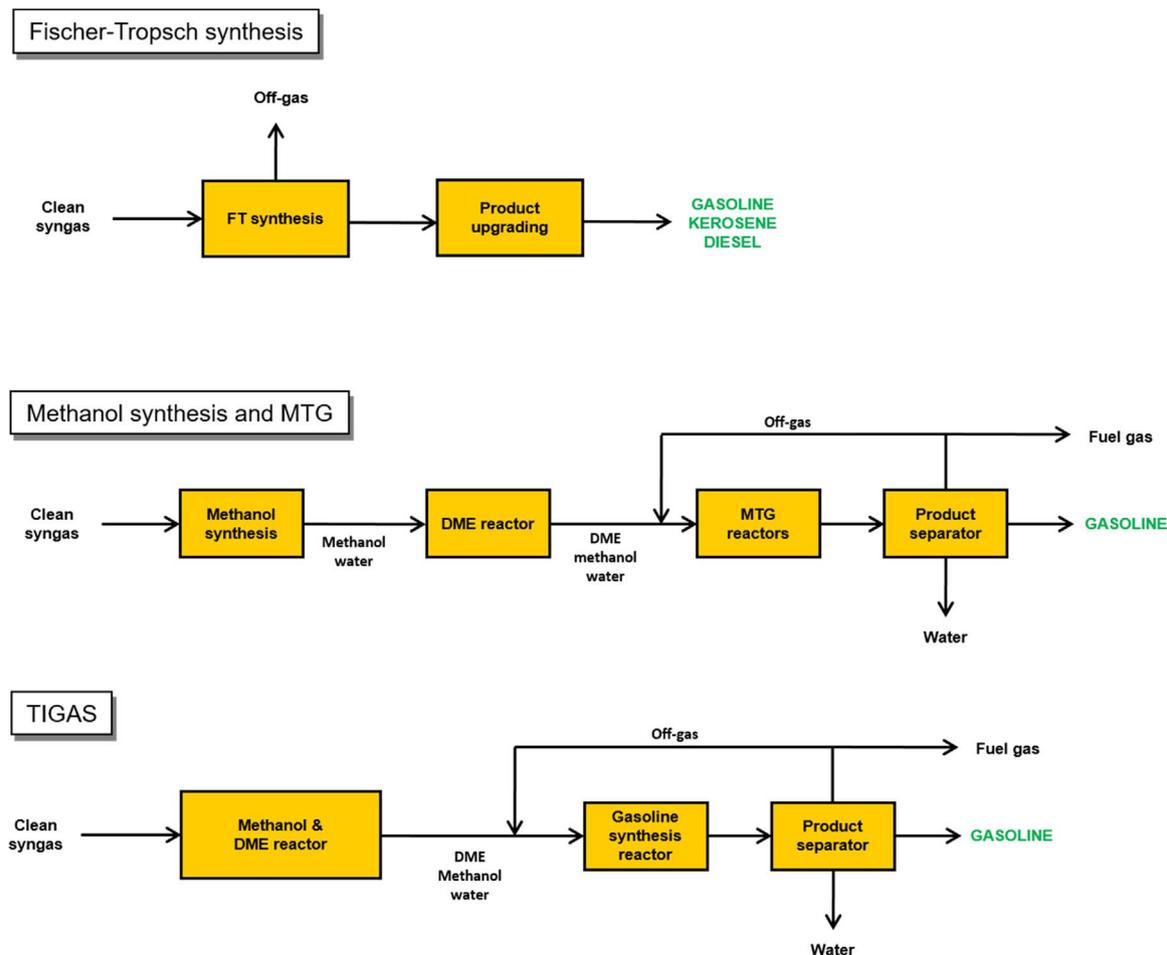


Fig. 2. Block flow diagrams of the fuel synthesis process concepts. FT: Fischer-Tropsch. DME: dimethyl ether. MTG: methanol-to-gasoline.

clean syngas [43,45].

For the EF gasification concepts, the syngas is fed to a direct water quench where it is cooled to the operating temperature of the WGS reactor (200 °C) [43]. The cooled syngas then passes through a bag filter to remove particulates. The H₂/CO ratio of the product gas from the EF gasifier is approximately 0.7, which is lower than the required ratio (H₂/CO = 2) for FT and methanol synthesis. As discussed in the previous section, the ratio is adjusted by using a WGS reactor which was modelled at equilibrium conditions and had an exit gas temperature of 340 °C [45]. For the Rectisol unit, the same conditions as the CFB gasifier were used.

3.6. Fuel synthesis

The clean syngas is the fed to the fuel synthesis section to be converted to liquid fuels by either FT synthesis, the MTG process or the TIGAS process, as discussed above. The block flow diagrams of FT synthesis, MTG and TIGAS process concepts selected in this study are shown in Fig. 2. The FT synthesis is specified as taking place over a cobalt-based catalyst at 230 °C and 25 bar [59]. The product distribution was estimated using the Anderson-Schulz-Flory (ASF) model with an alpha value of 0.85 which favours the production of middle distillates [27,43,83]. The single-pass fractional conversion of CO was adjusted to 80% [23]. Following Hamelinck [22], Furnsinn [27] and Swanson [43], all waxes are hydrocracked to middle distillate products (C₁₀-C₁₉). The product from the hydrocracking unit is isomerised to improve the cold flow properties and subsequently fractionated in a conventional distillation column. A product distribution of 60% diesel, 25% gasoline and 25% kerosene was assumed for the hydrocracking unit, as reported from Eilers et al. [84] for the Shell Middle Distillate Synthesis (SMDS) process.

For the MTG process, syngas from the gas clean-up passes first to the methanol synthesis reactor where it is converted to methanol at 50 bar and 250 °C [85,86]. The crude methanol is then fed to the MTG plant. Methanol synthesis catalysts have been reported to be extremely selective, with 99% or better selectivity to methanol [85]. According to LeBlanc et al. [85], methanol synthesis by-products are present in concentrations of less than 5000 ppm by weight and they are also reported to be converted by the downstream MTG process [25]. Therefore, given the high selectivity reported in the literature, no other by-products (e.g. higher alcohols, hydrocarbons) were included in the crude methanol product. Philips [25] reports 96 wt% methanol purity in the product, whereas Jones [26] assumes 93 wt%. In the ExxonMobil's MTG plant, methanol at 83 wt% was produced [85]. In this study, the methanol synthesis model was designed to produce methanol at 90% purity which is the approximate average of the reported literature values. The other 10% consists of other gas compounds (mainly H₂O, CO and CO₂) which were determined via elemental mass balances.

The simulation of the MTG process (process layout, operating conditions) was based on the ExxonMobil's MTG plant [67,87]. Methanol produced at the methanol synthesis plant is vaporised by heat exchange with MTG reactor effluent gases before it enters a dehydration reactor where a mixture of DME, methanol and water is produced at 404 °C. In order to estimate the elemental composition and product yield of the dehydration reactor effluent, the conversion of methanol to DME and H₂O was set to 77% [87]. The effluent from the DME reactor is combined with recycle gas from the product separator and enters the MTG reactor where it is converted at 415 °C and 21.2 bar to mainly hydrocarbons and water. The gasoline fraction in the product stream was set to 36 wt% of the methanol and DME input as reported by Yurchac [87] for the ExxonMobil MTG process. The conversion of methanol and DME was 100% [60,87,88] thus the product stream did not contain any methanol or DME.

The hot reactor effluent is cooled by heat exchange with the recycle gas from the vapour-liquid separator. It is then further cooled to about 200 °C before it passes to the vapour-liquid separator, where gas, liquid

gasoline and water separate. The off-gas from the product separator contains mostly low hydrocarbons (C₁-C₃), CO and CO₂. The water from the product separator contains trace amounts of oxygenated organic compounds and thus it requires treatment [60].

As discussed previously, the main difference between the TIGAS process and the MTG process is the absence of a discrete methanol synthesis step. In the TIGAS process, methanol and DME are synthesised in one reactor (oxygenate synthesis reactor) and they are then converted to gasoline in the gasoline synthesis reactor. As with the MTG process, the gasoline product is separated from gas and water in a vapour-liquid separator. The oxygenate synthesis is specified as taking place at 250 °C [23]. In order to estimate the mass balances of the oxygenate synthesis reactor effluent, the weight fraction of methanol and DME in the oxygenate product mix were set to 20% and 80%, respectively [61]. Other gas compounds (mainly H₂O and CO₂) were determined via elemental mass balances. The gasoline synthesis reactor of the TIGAS process used the same technology as the gasoline synthesis reactor of the MTG process [60]. Therefore, the model settings and operating conditions of the TIGAS gasoline reactor were the same as the MTG reactor.

4. Economic assessment

This section focuses on the costs of the six BTL process concepts evaluated in this study. One main purpose of the economic analysis was to identify the most promising BTL processes in terms of fuel production costs. Another important aim was to examine whether BTL plants can compete economically with conventional transport fuels plants. Table 4 summarises the general assumptions used for the economic assessment.

4.1. Capital costs

The total capital investment (TCI) of each BTL concept modelled is calculated using factorial estimation [89,90]. This is an established cost estimation method reported by Peters [89] whereby the TCI is calculated from the total purchased equipment cost (TPEC) by using ratios based on cost breakdowns for a solid-fluid processing plant, as shown in Table 5.

Firstly, the installed direct costs of the process sections described in Section 2 were calculated for each BTL process concept using reported costs from the literature. Table 6 shows the literature scales and installed direct costs which were used in this study. It is unusual to get a published cost estimate for the exact size of the plant that is being considered. Thus, when necessary, the installed direct costs were scaled by capacity as follows [89]:

$$IDC_{s,p}^{study} = IDC_{s,p}^{base} \times \left(\frac{S_{s,p}^{study}}{S_{s,p}^{base}} \right)^n \quad (1)$$

where $IDC_{s,p}^{base}$ and $S_{s,p}^{base}$ are the literature costs and capacity of the process section s for process p , $S_{s,p}^{study}$ is the equivalent capacity used in this study and n is the cost capacity exponent [89]. Where different studies give values for the $IDC_{s,p}^{base}$ of a process section, as is the case for gas conditioning – Rectisol, an average of the scaled costs is taken as $IDC_{s,p}^{PS}$. As there was no available information in the base study on the

Table 4
General economic parameters for the economic analysis.

Base year	2014
Plant life	20 years
Plant annual operating hours	8000
Loan interest rate	10%
Wood chips price	€55.54/dry tonne ^a

^a Initial value was \$70/dry tonne [77] converted and updated to 2014 EUR (€) using exchange rate and inflation rate from the Bank of England [78] and US Inflation Calculator [79], respectively.

Table 5
Calculation of total capital investment (TCI).

Cost parameter	Calculation method
Installed direct costs (IDC)	
Total purchased equipment cost (TPEC)	
Purchased equipment installation (PEI)	39% TPEC
Instrumentation and controls	26% TPEC
Piping	31% TPEC
Electrical systems	10% TPEC
Service facilities	55% TPEC
Non-installed direct costs (NIDC)	
Buildings	29% TPEC
Yard improvements	12% TPEC
Land	6% TPEC
Total direct costs (TDC)	IDC + NIDC
Total indirect costs (TIC)	
Engineering and supervision	32% TPEC
Construction expenses	34% TPEC
Contractor's fee and legal expenses	23% TPEC
Contingency	37% TPEC
Fixed capital investment (FCI)	TDC + TIC
Working capital (WC)	15% FCI [81]
Total capital investment (TCI)	FCI + WC

feedstock input for the fuel synthesis process section of MTG and TIGAS, the cost is scaled based on the plant's fuel output (t/d). Base fuel output is given in Table 5. For this study, the plant's energy output *EO* is calculated in GJ/h and can be converted to t/d using the following equation:

$$S_{s,p}^{study} = \frac{EO}{LHV_{fuel}/1000} \times 24 \tag{2}$$

where LHV_{fuel} is the liquid fuels LHV value for the given concept in Table 7.

The Chemical Engineering Plant Cost Index was used to update the costs to 2014 whenever a literature reported cost was from a previous year [89,90]. Most literature costs used in this study were installed costs. In the rare case when a reported cost was not an installed cost, it was converted to an installed cost using the factors shown in Table 5.

For a given process concept, the installed cost of the whole plant (*IDC*) was calculated as the sum of the installed costs of the individual process sections associated with the concept, as shown in Table 6:

$$IDC = (1 + \rho_{power}) \sum_{s,p \in concept} IDC_{s,p}^{study} \tag{3}$$

Table 6
Base scales and installed direct costs of major process sections used.

Process section	Process	Main items	Base capacity	Base cost (million US \$)	Base year	Study capacity	References
RSH & preparation	EF	Dryer, grinder	2000 dry t/d	21.3	2007	2016 dry t/d	Swanson et al. [31]
	CFB	Dryer	2000 dry t/d	19.3	2007	2016 dry t/d	Swanson et al. [31]
Air separation	EF	Distillation column	735 t/d O ₂	24.3	2007	1213 t/d O ₂	Swanson et al. [31]
	CFB	Distillation column	735 t/d O ₂	24.3	2007	881 t/d O ₂	Swanson et al. [31]
Gasification & gas cleaning	EF	EF gasifier, lockhopper, Water quench	2000 dry t/d	63.4	2007	2016 dry t/d	Swanson et al. [31] Reed [82]
	CFB	CFB gasifier, lockhopper, tar cracker, syngas cooler	4536 dry t/d	112.9	2009	2016 dry t/d	Larson et al. [12]
Gas conditioning - WGS	EF	WGS reactor	1650 dry t/d	2	1991	2016 dry t/d	Williams et al. [83]
Gas conditioning - Rectisol	EF & CFB	Absorber, stripper	1800 dry t/d	14.5	1992	2016 dry t/d	WVU [84]
			4536 dry t/d	44	2009		Larson et al. [12]
			1650 dry t/d	38	1991	2016 dry t/d	Williams et al. [83]
Fuel synthesis (MeOH)	MTG	MeOH reactor	2000 dry t/d	42.2	2007	2016 dry t/d	Swanson et al. [31]
Fuel synthesis	MTG & TIGAS	FT reactor	1735 t/d	83.5	1982	Fuel energy output dependent	Grace et al. [85]
		DME reactor, MTG reactor, refining	1149 t/d gasoline	117	1988		Bridgwater et al. [86]
			2000 dry t/d	29.5	2007	2016 dry t/d	Swanson et al. [31]
Refining	FT	Hydrocracking	14% of other installed equipment costs				[12,16,31]
Power generation	All	Steam turbine					

where the summation is all stages and processes associated with the concept and ρ_{power} is the contribution of power generation to *IDC* (14%).

The installed direct costs were then used to calculate the total capital investment using the factors and methodology given in Table 5. Note that costs are estimated as a percentage of purchased equipment cost (*TPEC*) and the costs calculated from literature above are installed costs (*IDC*), so first *TPEC* was calculated by dividing *IDC* by the sum of the factors associated with all the installed direct costs. Knowing *TPEC* allows us to use the factors shown in Table 5 in order to determine the fixed capital investment:

$$FCI = \frac{IDC}{2.61} \times (4.28 + \rho_{land}) \tag{4}$$

where ρ_{land} is the contribution to non-installed direct costs of land (6%) which is left as a parameter for use in Section 5. Finally, the TCI was calculated for each process concept as the sum of the FCI and working capital (*WC*):

4.2. Production costs

The total annual costs consist of annual capital repayments, as well as operating and maintenance costs (e.g. biomass costs, utilities, fixed charges, labour costs). For each BTL process concept, the fuel production costs are calculated as follows:

$$Fuel\ production\ costs = \frac{ACR + TOMC}{OH \times EO} \tag{5}$$

where *ACR* is the annual capital repayment, *TOMC* is the annual total operating and maintenance costs, *OH* are the plant's annual operating hours and *EO* is the fuel energy output (GJ/h).

The price inflation of equipment and raw materials is not considered for the ease of comparison between the evaluated BTL concepts. Similarly, government subsidies and by-product revenues are excluded from the economic analysis.

4.2.1. Annual capital repayment

This cost calculation method amortizes the installed capital investment over the anticipated life of the plant at a given interest rate. The annual capital repayment is the money required to pay back the loan on capital which is required to set up the plant. It is calculated from the following equation [91]:

Table 7
Mass and energy balances.

		EF-FT	EF-MTG	EF-TIG	CFB-FT	CFB-MTG	CFB-TIG
Biomass							
Mass flow (wet)	kg/h	120,000	120,000	120,000	120,000	120,000	120,000
Mass flow (dry)	kg/h	84,000	84,000	84,000	84,000	84,000	84,000
LHV ^a	kJ/kg	19,897	19,897	19,897	19,897	19,897	19,897
Power input ^a	MW	459.6	459.6	459.6	459.6	459.6	459.6
Liquid fuels							
Mass flow	kg/h	16,958	14,810	15,831	17,930	16,980	17,541
LHV	kJ/kg	43,917	42,308	42,338	43,917	42,307	42,312
Fuel energy output	GJ/h	744.84	626.76	670.32	787.32	718.2	742.32
Power output	MW	206.9	174.1	186.2	218.7	199.5	206.2
Yield & efficiency							
Fuel energy efficiency	%	45	37.9	40.5	47.6	43.4	44.9
Fuel mass yield	%	20.4	17.8	19	21.6	20.4	21

^a Values are given on a dry ash-free basis.

$$ACR = TCI \cdot \frac{r \cdot (1+r)^N}{(1+r)^N - 1} \quad (6)$$

where ACR is the annual capital repayment, TCI the total capital investment, r the interest rate and N the plant life.

4.2.2. Operating and maintenance costs

There appears to be limited consensus in the literature as to the staffing level, utilities costs and requirements for large-scale BTL plants. To deal with this problem operating and maintenance (O&M) costs can be taken instead as a percentage of the fixed capital investment. Van Vliet et al. [24] assume that the O&M costs (labour, maintenance, raw materials, waste disposal, utilities) are 4% of the fixed capital investment (feedstock costs and fixed charges are not included). Hamelinck et al. [92], as well as Larson et al. [23] also use 4% FCI for the O&M costs of large-scale BTL plants. Therefore, in this study, O&M costs were assumed to be 4% of the fixed capital investment. Adding 3% of FCI as fixed charges (insurance, taxes) [89,90], the total O&M costs, excluding biomass costs, were 7% of FCI.

The price of wood chips was €55.54/dry tonne (see also Table 4) and this included growing, harvesting and storage, as well as transporting biomass to the gasification plant [13]. The cost of biomass can vary significantly and it mainly depends on the biomass yield and the land available for cultivation [39]. Other studies have shown that the biomass feedstock cost is an important element of the biofuel production costs [13,16].

Combining the above, the total operating and maintenance costs are given by:

$$TOMC = (\rho_{OM} + \rho_{FC}) \times FCI + PC \times OH \times FdC \quad (7)$$

where ρ_{FC} and ρ_{OM} are the proportions of FCI used to estimate respectively the fixed costs and the other O&M costs (3% and 4%, see above). PC is the plant capacity (dry tonnes/h), OH are the plant's yearly operating hours and FdC is the feedstock cost (€/dry tonne).

5. Results

5.1. Energy efficiency

The fuel energy efficiency is widely used to assess the technical performance of BTL plants [16,23]. It is a measure as to which extent the energy in the biomass feedstock remains in the hydrocarbon fuel products. In this study, it is defined as the ratio between the total energy in the hydrocarbon fuels and that in the biomass feedstock:

$$\eta_{plant} = \frac{\dot{M}_{fuels} \cdot LHV_{fuels}}{\dot{M}_{biomass} \cdot LHV_{biomass}} \quad (8)$$

where \dot{M}_{fuels} is the mass flow (kg/h) of hydrocarbon fuels, LHV_{fuels} is the

lower heating value (kJ/kg) of hydrocarbon fuels, $\dot{M}_{biomass}$ is the mass flow input (kg/h) of the dry ash-free biomass and $LHV_{biomass}$ is the lower heating value (kJ/kg) of the dry ash-free biomass feed. The energy efficiency as defined above was calculated for each selected BTL plant concept in order to compare the different BTL concepts in terms of performance.

The mass and energy balances as well as the fuel energy efficiencies for all process concepts are presented in Table 7. Looking at the corresponding efficiencies shown in Table 7 it becomes clear that the FT concepts perform better than the TIGAS and MTG concepts. Specifically, the CFB-FT concept shows the highest fuel energy efficiency at 47.6%. This is due to the additional synthesis steps required in TIGAS and especially in the MTG process to produce liquid hydrocarbon fuels. The TIGAS process requires initially the synthesis of methanol and DME in the oxygenate synthesis reactor and then the synthesis of gasoline in another reactor. The MTG process includes three subsequent synthesis steps: methanol synthesis, oxygenate synthesis and gasoline synthesis. As 100% conversion to the desired products is not achieved in any individual synthesis step, each additional synthesis step results in lower mass yields and thus lower energy conversion efficiencies. It can also be seen that the CFB concepts deliver higher fuel energy efficiencies by 5.7–14.7% than the EF concepts with the same fuel production technology as the H₂/CO molar ratio of the clean syngas in the CFB concepts is higher and closer to the optimum ratio for fuel synthesis.

This study's results on the energy conversion efficiency of BTL plant concepts are consistent with those from other modelling studies which range from 34% to 52% (LHV) [13,16,23]. As a large-scale BTL plant has yet to be built the results of this study cannot be compared with a real industrial BTL plant. Bridgwater [8] reports that mass yields of BTL plants based on biomass gasification range from 14.9% to 23.5% on dry basis. The mass yield results of this study are well within this range.

5.2. Costs

5.2.1. Capital costs

Fig. 3 shows the breakdown of capital costs by process area and the total capital investment for all six BTL plant concepts. The cost data used for Fig. 3 is shown in Supplementary, Table S1. The capital costs of large-scale BTL plants range from €397–505 million. All the MTG based concepts have higher capital costs by up to 27% than the equivalent FT and TIGAS based concepts. This difference is easy to understand since the MTG process includes the additional conversion step of methanol synthesis. A significant portion of the methanol synthesis capital cost is the syngas compression to 50 bar which is the operating pressure of methanol synthesis. According to Swanson et al. [43], compressors have high purchase costs and can make up of approximately 18% of the purchased equipment costs of BTL plants. The additional compression step required for the MTG process would result in higher compression

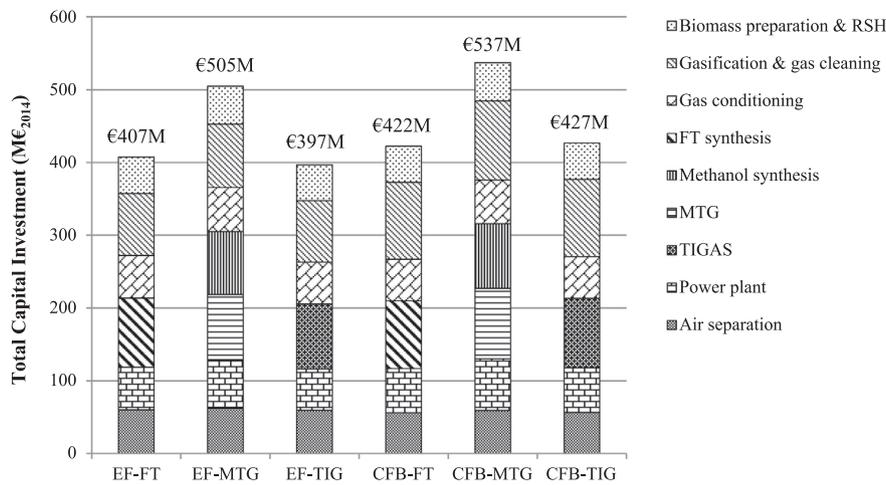


Fig. 3. Total capital investment of the evaluated BTL process designs.

equipment requirements and thus higher capital costs than FT and TIGAS which operate at a similar pressure to the EF and CFB gasifiers.

The major process areas of capital investment for all BTL concepts are the gasification and gas clean-up area as well as the fuel synthesis area. This is in line with other techno-economic studies of BTL plants [13,16]. The BTL concepts based on the CFB gasification technology have higher gasification and gas clean-up costs than the EF based concepts due to the need for a tar cracker and the additional heat exchange equipment required for syngas cooling. As discussed previously, syngas cooling for the EF concepts was done by a direct water quench thus a heat exchanger was not employed.

This study's results on the capital costs of BTL plant concepts are consistent with those from other BTL techno-economic modelling studies of similar scale which range from €340 to €499 million (reported costs were adjusted to 2014) [13,16]. However, cost comparisons with other studies should be made with caution due to the different financing assumptions of each study.

5.2.2. Operating and maintenance costs

Fig. 4 shows the annual operating and maintenance costs which include biomass costs and capital dependent operating costs, as discussed in Section 4.2.2. The cost data used for Fig. 4 is shown in Supplementary, Table S2. The total O&M costs of large-scale BTL plants range from €62 to 68 million. Biomass costs are the largest contributor to O&M costs and represent essentially more than 50% of the annual O&M costs for all BTL concepts. Once again the MTG concepts have higher costs than the FT and TIGAS concepts. This was expected since most of the O&M costs are a fraction of the FCI which is higher for the MTG

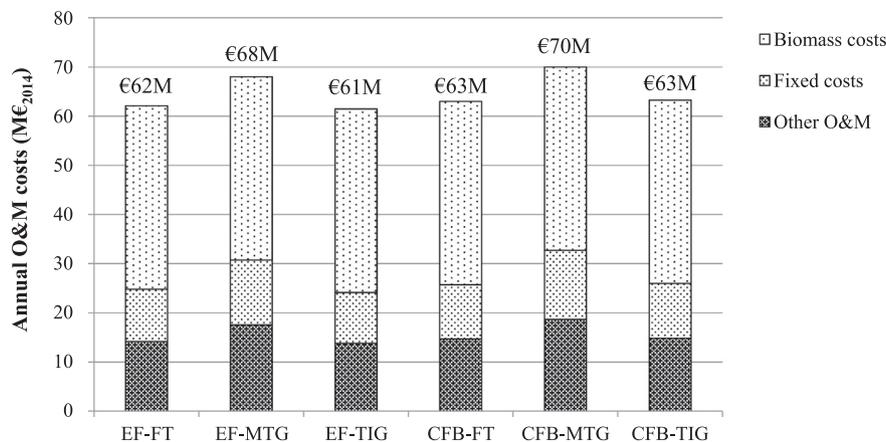


Fig. 4. Annual operating and maintenance costs of the evaluated BTL process designs.

Table 8

Liquid fuel production costs for the evaluated BTL process designs.

	EF-FT	EF-MTG	EF-TIG	CFB-FT	CFB-MTG	CFB-TIG
Capital (€GJ ⁻¹)	8.03	11.83	8.69	7.88	10.98	8.44
O&M (€GJ ⁻¹)	4.16	6.13	4.50	4.08	5.69	4.37
Biomass (€GJ ⁻¹)	6.26	7.45	6.96	5.92	6.49	6.29
Total (€GJ⁻¹)	18.46	25.41	20.16	17.88	23.17	19.10

concepts.

5.2.3. Production costs

The production costs per GJ of gasoline and diesel for all BTL cases considered are presented in Table 8, along with the contribution of capital costs (as capital annuity), O&M expenditure and biomass cost. The calculated production costs only include the necessary expenditure to manufacture 1 GJ (LHV) of fuel, i.e. they do not include tax, duties, profits, marketing and distribution costs.

It can be seen that the BTL plants based on FT synthesis have lower production costs than the TIGAS and MTG based BTL plants. More specifically, the CFB-FT concept gives the lowest production costs of liquid hydrocarbon fuels at €17.88 per GJ. It is closely followed by the EF-FT concept at €18.46 per GJ. Even though EF-FT results in lower capital and O&M costs than CFB-FT (see Figs. 3 and 4), the latter has lower production costs due to its higher fuel production. From this it is clear that the plant fuel output is a very important element of the production costs thus its effect is investigated in the sensitivity and uncertainty analysis sections later in the paper.

Table 8 also shows that capital expenditure is the most important contributor to the fuel production costs as it represents 43–47% of the total production costs. Biomass costs also have a significant effect representing approximately 28–35% of the fuel production costs of the BTL process designs. As facilities get larger, the feedstock contribution increases since economies of scale reduce the contribution from capital costs. The fuel production costs estimated in this work are consistent with those from other BTL techno-economic modelling studies that considered similar plant capacities and range from €13 to €30 per GJ (reported costs were adjusted to 2014) [13,16].

5.2.4. Comparison with market price of conventional transport fuels

The 2014 refinery gate price (i.e. excludes tax, duty, profits, marketing and distribution costs) of conventional diesel and gasoline was €16.2 and €16.6 per GJ, respectively [21]. The reported fuel prices are given in £/l and they were converted to €/GJ by using the Bank of England 2014 exchange rate [93] as well as a volumetric energy density of 34 MJ/l (LHV) and 32 MJ/l (LHV) for diesel and gasoline, respectively [94]. In comparison, the production cost of CFB-FT (diesel as main product) is 7.7% higher than the market price of petroleum-derived diesel. For the best gasoline-based concept (CFB-TIG), the production cost is 17.6% higher than the reference gasoline price. The other concepts will require further development to reduce costs before they can be economic but viability by these two routes is a likely prospect.

For second generation transport fuels to become more competitive with fossil-derived fuels, increases in fossil fuel prices and subsidies or changes in legislation, such as higher carbon taxes are required. For the CFB-FT concept, which was the most economic BTL system, a subsidy of approximately €12/tonne of dried wood would be necessary to meet the conventional fuel prices. However, promotional measures, such as biomass and biofuel subsidies (e.g. UK's Renewable Transport Fuels Obligation scheme) are usually aimed at small or medium-sized power plants and it is therefore questionable whether they can be applied for large scale BTL installations. Government subsidies would mainly depend on the political climate and thus their level is still unknown. Many government programmes have limited budget or are tied to certain technologies, thus a project of this size might not be subsidised. For example, in the UK, renewable electricity was not originally taxed under the Climate Change Levy scheme; however, in August 2015 the UK government decided to remove this exemption due to budget restrictions. Although bioenergy subsidy schemes are valuable opportunities for BTL plants in the short term, it would be far better to find markets (e.g. chemicals) that could be penetrated without any subsidies.

5.3. Sensitivity analysis

The evaluations thus far have taken the results of the models at face value, therefore sensitivities and uncertainties in the models have not yet been considered. This section presents a sensitivity analysis study to investigate the effect of parameters variations on the production costs results.

In the previous sections, some parameters were identified as having an important effect on the production costs of liquid hydrocarbon fuels. These parameters were selected for the sensitivity analysis and were: fuel output (kg/h), capital costs (i.e. TCI) and biomass cost. Other parameters included: O&M costs (as a percentage of FCI, see Section 4.2.2), loan interest rate, plant operating hours and plant life as these were identified by other studies for also significantly affecting production costs [43,95]. The sensitivity analysis was carried out by changing each parameter in turn by $\pm 30\%$ of its base case value, except for the plant operating hours which were changed by $\pm 9.5\%$ since they cannot exceed the maximum hours per year. The sensitivity analysis results of the CFB-FT concept which has the lowest production costs are shown in Fig. 5. The bars show deviations from the original

values of the model parameters with longer bars indicating a higher degree of sensitivity to a particular parameter. Similar diagrams for the other five BTL process designs are available in Supplementary, Figs. S1–S5.

Fig. 5 shows that the fuel output has the greatest effect on production costs which can drop to €13.75 per GJ (or 23% below the base case cost) when the fuel output is increased by 30%. The performance of CFB gasification and FT synthesis reactors is established in general; however, there is limited experience of operation of these reactors for biofuel production. This increases the uncertainty of the overall results. The sensitivity of the model to the biofuel product output suggests that improving the performance of CFB gasification and FT synthesis technologies should be an early priority.

The capital investment is the next most sensitive parameter. The fuel production costs of CFB-FT can be reduced by 13% if the total capital is decreased by 30%. Estimate uncertainties of 30% for capital costs are typical in BTL studies which are based on literature cost data and the factorial estimation method [19,23]. Significantly increased accuracy can only be achieved through acquiring capital cost data from a commercial BTL plant which is not currently available.

The biomass cost is the third most sensitive parameter; however, this cost can vary considerably in different locations and certainly well outside the 30% limit tested here. Production costs increase to €20.24 per GJ if the wood chips price rises by 30% (€72.2 per dry tonne). On the other hand, if the biomass price drops by 30% (€38.9 per dry tonne), the production costs will be 3% lower than the market price of conventional fuels making this route economically viable. To avoid increases in the biomass price, the conclusion of a long term biomass supply contract with fixed quantities and prices should be sought with an agricultural or forest management company. If it is decided that several companies should supply biomass during a large scale BTL plant project, a joint delivery commitment should be agreed. Finally, the biofuel production costs are less sensitive to the loan interest, plant operating hours, O&M costs and project's lifetime.

5.4. Uncertainty analysis

Uncertainty in the model's output (i.e. production costs) results mainly from uncertain estimates of various model parameter values. Many of these values were taken from existing BTL techno-economic studies with their own estimates and assumptions and thus they depend on the accuracy and reliability of each published study. Uncertainty in the model's output can also result from errors in the simulation model structure compared to a real system, and approximations made by numerical methods employed in the simulation. Process simulation models are always simplifications of real processes and, hence, 'imprecision' can result. Increasing the model complexity to more closely represent the complexity of the real system may not only significantly add to the time and cost of data collection, but may also introduce even more parameters, and thus even more potential sources of uncertainty in the model's output.

In this study, the uncertainty analysis employed the Monte Carlo method which is commonly used for uncertainty analysis studies [31]. This method is based on random sampling from the probability distributions of a model's input parameters and repeated runs to estimate the probability distribution of the model's output. Each selected technical and cost parameter of the BTL cost models was assigned a range of values and a probability distribution derived from the literature and industry experts. A Monte Carlo simulation of the selected input parameters to account for uncertainties was then implemented in the programming language C++ in which values for each of the uncertain parameters were drawn from their assumed distributions and used to estimate the production costs (Eq. (5)) according to the methodology described in Section 4 resulting in a single production cost sample for a given process concept. One million such samples were drawn for each concept and normalised histograms were then used to approximate the

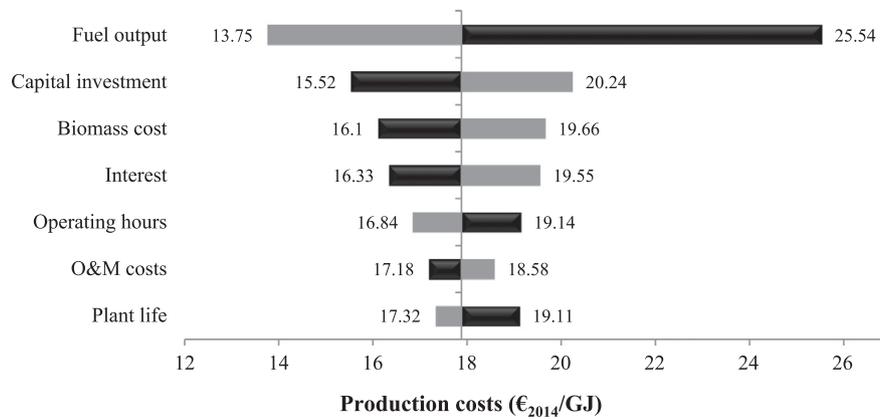


Fig. 5. Sensitivity of fuel production costs of CFB-FT to variations of selected technical and economic parameters (all parameters are varied by + 30% (grey) and - 30% (black), apart from the plant operating hours which are varied by ± 9.5%). The vertical line in the graphs represents the production cost of the base case for CFB-FT.

probability density functions of the biofuel production costs for each concept.

The selected uncertain parameters together with their value range are shown in Table 9. The choice and the value range of the uncertain input parameters of the cost model emerge from the sensitivity analysis (see Section 5.3), the literature and experts’ opinion. As we have more data supporting the base assumptions/estimations than we do supporting the ranges, we take all parameter values to be normally distributed with mean equal to the base assumptions. In the absence of data from which to calculate the standard deviation (σ) of the distributions, we follow the three-sigma rule to ensure that “almost all” (>99.7%) of the distribution lies in the specified range by setting σ for each distribution such that the furthest part of the range from the mean lies 3σ away. For example, the base estimation of the power plant installed cost ratio is 14% (0.14) and the furthest part of the range from the base estimation is 11% (0.11). Therefore, $\sigma = \frac{0.14 - 0.11}{3} = 0.01$ so $\rho_{power} \sim N(0.14, 0.01^2)$. Any values which lie outside of the ranges described in Table 9 are discarded and regenerated.

The cumulative probability distributions of fuel production costs of the CFB concepts and the EF concepts are shown in Figs. 6 and 7, respectively. They show a significant range of cost values with non-negligible probability. For example, the CFB-FT concept which has the lowest production costs has a 90% confidence interval for production cost of €15.66–22.13 per GJ, compared to the deterministic estimate of €17.88 per GJ. In other words, the actual value has a 90% chance to be within the range of €15.66 to €22.13 per GJ. This illustrates the potential risks and rewards associated with this process concept. On one hand, the price of conventional transport fuels (average price: €16.4 per GJ, see Section 5.2.4) lies within the 90% confidence interval meaning that there is a realistic chance of the concept being economically viable

Table 9
Selected uncertain input parameters and their value range.

Parameter	Base assumption/ estimation	Range	Source
Biomass cost (FdC)	€55.54/dry tonne	± 50%	[92]
Installed costs ($IDC_{s,p}^{base}$)	See Table 6	± 30%	[12,31]
Cost capacity exponent (n)	0.65	0.6–0.7	[80]
O&M costs, excluding fixed costs (ρ_{OM})	4% FCI	3–5%	[8,12]
Fixed costs (ρ_{FC})	3% FCI	2–4%	[80,81]
Interest rate (r)	10%	8–15%	[92]
Land (ρ_{land})	6% TPEC	4–8%	[80]
Power plant installed cost (ρ_{power})	14% other installed costs	11–16%	[12,16,31]
Plant operating hours (OH)	8000 h/year	7008–8322	[16,32]
Fuel energy output (EO)	See Table 7	± 20%	[92]

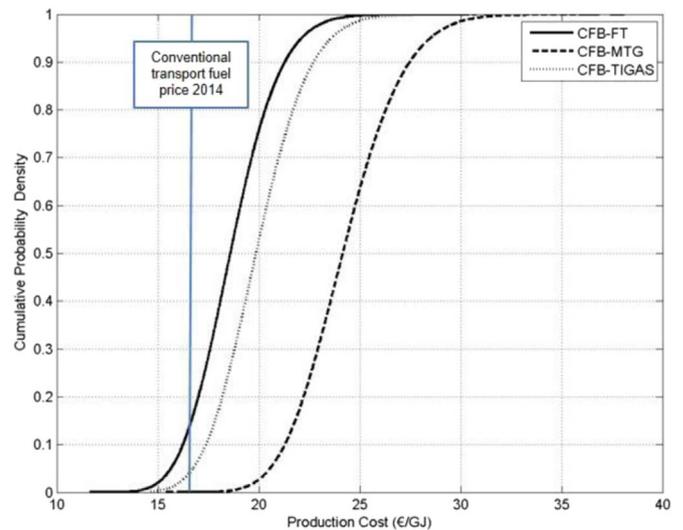


Fig. 6. Cumulative probability of biofuel production costs of the CFB concepts.

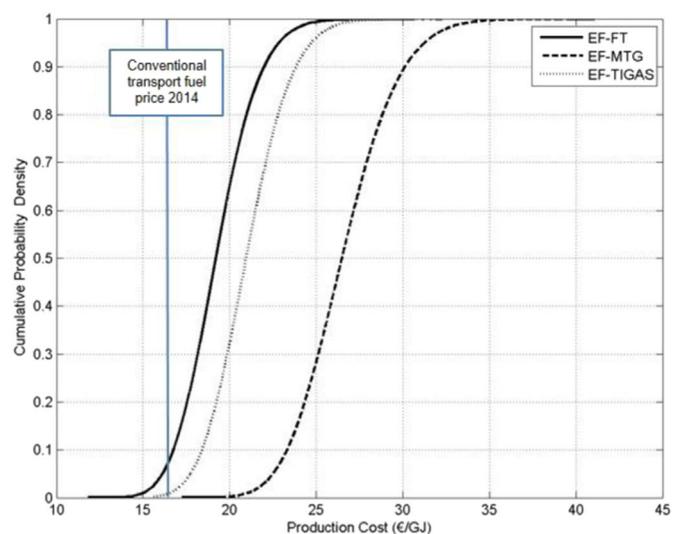


Fig. 7. Cumulative probability of biofuel production costs of the EF concepts.

in comparison to conventional fuels. On the other hand, a price at the upper end of that range would be far from economic. As discussed previously, risk-reward information of this type is very useful to potential investors and may make an investment in this technology more

attractive than a deterministic estimate which is above the price of conventional fuels and carries no information about risk. These results can also be used directly to estimate the probability of one of the process concepts examined meeting or bettering the price of conventional fuels. The two most promising concepts in this regard are CFB-FT and EF-FT which have, respectively, a 14% chance and an 8.4% chance to meet or better the conventional transport fuel price.

Another area in which these results could be valuable is when setting policy to encourage investment in BTL technologies, for instance by offering tax incentives [96,97]. Knowledge of the probability distributions of the production costs allow this to be done in a principled way. For example, they would allow incentives to be set in such a way that a given process concept had a 50% chance of being economically competitive. In 2014, the conventional transport fuel tax (duty and VAT) rate in the UK was 60% [21]; therefore, the conventional fuel price including tax was approximately €41.5 per GJ. If biofuels from the CFB-FT concept were to have 50% chance (€18.6 per GJ in Fig. 6) to meet this price then a tax rate of approximately 55% would be required. Therefore, if the government fuel tax rate was reduced by 8%, biofuels through the CFB-FT route could be competitive with conventional transport fuels. Table 10 shows the tax rate reduction required for the BTL designs to meet the conventional fuel prices with 50% probability. The most expensive option (EF-MTG) would need approximately five times higher tax rate reduction compared to CFB-FT which has the lowest production costs of all process designs.

Finally, the results show that the deterministic estimates of production cost systematically underestimate the likely cost. In all six process concepts, the deterministic estimates of costs shown in Table 8 are lower than the median result from the uncertainty analysis (Table 10). Although the deterministic estimates may be the single most likely cost figures, these results show that they have less than a 50% chance of being achieved. For example, Fig. 6 shows that there is a 35% chance that the production cost of the CFB-FT concept will meet or better the deterministic estimate of €17.88 per GJ.

6. Conclusions

This work has examined the technical and economic feasibility of six Biomass-to-Liquid (BTL) process configurations for the production of liquid transport fuels based on entrained flow (EF) and circulating fluidised bed (CFB) gasification of wood biomass. Fuel synthesis technologies included Fischer-Tropsch synthesis, methanol synthesis followed by the Methanol-to-Gasoline (MTG) process and the Topsoe integrated gasoline (TIGAS) synthesis. Detailed designs were developed with the process simulation software IPSEpro to determine the technical and economic potential of the selected process configurations and identify the concept with the lowest overall costs. An uncertainty analysis based on the Monte Carlo statistical method was also carried out to examine the effect of uncertainties of the model input parameters and estimate the probability distributions of production costs.

The overall energy efficiency and production costs of the BTL concepts evaluated range from 37.9% to 47.6% LHV and €17.88–25.41 per GJ of produced fuels, respectively. Fischer-Tropsch synthesis seems to be the most promising fuel synthesis technology for commercial

Table 10

Tax rate reduction (%) required for biofuels to meet the conventional fuel prices with 50% probability. The median is the 50% probability point in Figs. 6 and 7.

	Median (€GJ ⁻¹)	Required tax rate (%)	Tax rate reduction (%)
CFB-FT	18.61	55.2	8.0
CFB-MTG	24.19	43.6	29.9
CFB-TIG	19.88	53.6	13.7
EF-FT	19.2	53.7	10.5
EF-MTG	26.51	38.2	38.6
EF-TIG	20.97	51.1	17.8

production of liquid fuels via biomass gasification since it achieved higher efficiencies and lower costs compared to TIGAS and MTG. This is due to the additional synthesis steps required in TIGAS and especially in the MTG process as each synthesis step adds to the overall costs, results in lower mass yields and thus lower overall energy conversion efficiencies.

The fuel synthesis concepts that incorporate circulating fluidised bed gasification technology have higher fuel energy efficiencies and lower production costs than the equivalent concepts based on entrained flow gasification by 5.7–14.7% and 3.2–9.7%, respectively. Even though the BTL concepts based on the CFB gasification technology have higher capital costs than the equivalent EF based concepts, the higher carbon conversion for the CFB gasifier, mainly due to the inclusion of a tar cracker, has a compensating effect and results in lower production costs. These results and the fact that there has been limited experience with entrained flow gasification of biomass so far, suggest that the circulating fluidised bed gasification technology is more promising in the short-term for large-scale production of second generation transport biofuels.

The resulting production costs for biomass-derived fuels via CFB gasification are 7.7%, 17.6% and 42.7% higher than the current market fuel price for the FT synthesis, TIGAS and MTG cases, respectively. The sensitivity analysis reveals that the fuel production costs are mainly influenced by variations in fuel product output (i.e. conversion efficiency), capital investment and biomass costs. This emphasises the importance of optimising current BTL technology, as well as the significance of long term biomass supply contracts with fixed quantities and prices. For the CFB-FT concept, which was the most economic BTL system, a subsidy of approximately €12/tonne of dried wood would be necessary to meet the conventional fuel prices.

Based on the initial cost estimates for this concept, the uncertainty analysis shows that, for the most promising concept (CFB-FT), there is a 14% probability that biofuel production costs will meet the price of conventional fuels without any subsidies. Additionally, biofuels via this route have a 50% chance to be competitive with conventional fuels if the government fuel tax rate was reduced by only 8%. The uncertainty analysis also indicates that deterministic estimates or sensitivity analyses of production costs may systematically underestimate the production cost of biofuels as they do not account for the effect of simultaneous variations of parameters.

While the evaluation showed that none of the BTL systems are currently competitive on price with conventional large scale fossil fuel plants (which enjoy the benefits of low feedstock costs and significant economies of scale), large scale biofuel production can be made competitive through a combination of moderate subsidies and tax reduction. Subsidy schemes are now in place in many countries in Europe and the UK but rely on public and political support for their long term implementation. Additionally, environmental taxes, such as greenhouse gas penalties would enhance the competitiveness of biofuels.

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Appendix A. Supplementary material

Supplementary data associated with this article can be found in the online version at <http://dx.doi.org/10.1016/j.rser.2018.02.023>.

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