

Paper IV

Ilkka Hannula

**Co-production of synthetic fuels and district heat
from biomass residues, carbon dioxide and electricity:
Performance and cost analysis**

Biomass and Bioenergy, 2015
Volume 74, Pages 26-46.

Copyright 2015 Elsevier
Reprinted with permission from the publisher

Available online at www.sciencedirect.com

ScienceDirect

<http://www.elsevier.com/locate/biombioe>

Co-production of synthetic fuels and district heat from biomass residues, carbon dioxide and electricity: Performance and cost analysis



Ilkka Hannula*

Technical Research Centre of Finland, P.O. Box 1000, FI-02044 VTT, Finland

ARTICLE INFO

Article history:

Received 13 August 2014

Received in revised form

14 December 2014

Accepted 6 January 2015

Available online

Keywords:

Biomass residues

Gasification

Power-to-fuels

Carbon dioxide

Synthetic fuels

District heating

ABSTRACT

Large-scale systems suitable for the production of synthetic natural gas (SNG), methanol or gasoline (MTG) are examined using a self-consistent design, simulation and cost analysis framework. Three basic production routes are considered: (1) production from biomass via gasification; (2) from carbon dioxide and electricity via water electrolysis; (3) from biomass and electricity via hybrid process combining elements from routes (1) and (2). Process designs are developed based on technologies that are either commercially available or successfully demonstrated at precommercial scale. The prospective economics of future facilities coproducing fuels and district heat are evaluated from the perspective of a synthetic fuel producer. The levelised production costs range from 18–37 €/GJ for natural gas, 21–40 €/GJ for methanol and 23–48 €/GJ for gasoline, depending on the production route. For a given end-product, the lowest costs are associated with thermochemical plant configurations, followed by hybrid and electrochemical plants.

© 2015 Elsevier Ltd. All rights reserved.

1. Background and scope

Deep reductions in anthropogenic emissions are required to stabilise the levels of atmospheric carbon dioxide (CO₂) [1]. As transportation and power generation are the two largest sources of global CO₂ emissions, they are also the most critical sectors of the economy where cuts need to take place [2].

In the power sector, near-term solutions for CO₂ management include photovoltaics, wind power, biopower, nuclear power and carbon capture and storage (CCS) technologies. Most of these options are ready for large-scale deployment and capable of inducing deep emissions cuts [3].

In the transportation sector, emissions can mainly be reduced by improvements in efficiency and change in vehicle fuel. However, most of the alternative fuel options (e.g. starch-based ethanol, biogas and electricity) require modifications to the current vehicle fleet and/or fuel distribution infrastructure, which severely limits the near-term potential for emissions cuts from the sector. For the medium-term, synthetically manufactured fuels (synfuels) are attracting attention as a way to produce alternative fuels that are compatible with the existing transportation infrastructure [4].

Technology for the production of synthetic fuels from fossil feedstocks, such as coal or shale, has existed for almost a century. However, when coal is used as feedstock, the resulting net greenhouse gas (GHG) emissions are about

* Tel.: +358 40 838 0960.

E-mail address: ilkka.hannula@vtt.fi.<http://dx.doi.org/10.1016/j.biombioe.2015.01.006>

0961-9534/© 2015 Elsevier Ltd. All rights reserved.

double of those from petroleum fuels [5]. It is possible to cut down part of these emissions with capture and storage of the byproduct CO₂, but the net GHG emissions would still be reduced only to levels comparable to those from petroleum fuels [6].

Switching partly or completely from fossil feedstocks to biomass (plant matter) is a frequently proposed method for further decarbonisation of synthetic fuel production [6–9]. Unfortunately, all commercial scale synfuels plants to date have been operated with fossil feedstocks and redesign of some key parts of the process is required to make the switch to biomass possible. Currently, a lot of RD&D work is ongoing to commercialise such technology [10].

Another solution would be to produce synthetic fuels directly from carbon dioxide and renewable electricity with a process referred to here as 'power-to-fuels' (P2F). This process begins with splitting water (H₂O) into hydrogen (H₂) and oxygen (O₂) with electricity. The produced hydrogen is then synthesised with co-feed CO₂ to form hydrocarbons or alcohols. The hydrogen is thus stored chemically as conventional liquid or gaseous fuel that can be consumed at a chosen time and place within the existing infrastructure. In this sense, the power-to-fuels concept elegantly solves problems of distribution and storage that normally impede energy concepts based on hydrogen production. However, the present use of CO₂ as chemical feedstock is limited to few industrial processes, although commercial projects based on hydrogenation of CO₂ to synthetic fuels are already emerging [11].

Yet another solution would be to combine the above-described processes together into a hybrid process that exploits biomass gasification to produce CO but uses renewable electricity to make up for the hydrogen deficit in the produced syngas. This combination would not require any new equipment to be developed while it also provides a solution to the lack of large-scale catalyst systems capable of direct hydrogenation of carbon dioxide.

The objective of this paper is to investigate the production of synthetic¹ fuels from biomass residues, CO₂ and electricity and their potential role in decarbonisation of the transportation fuel pool. A unified analytical framework is employed to systematically analyse and compare different plant configurations based on their mass and energy balances calculated with ASPEN Plus® (Aspen) process simulation software. The overall economics are evaluated under alternative feedstock price assumptions in terms of euros (€) per gigajoule (GJ), based on an underlying component-level capital cost estimates.

2. Plant configurations

All the analysed plant configurations feature two basic parts: synthesis gas production (endothermic) and synthesis gas conversion (exothermic). Energy integration between these two parts is possible to a certain degree via steam. The configurations considered here illustrate three basic alternatives:

- Production from biomass via gasification;
- Production from carbon dioxide and electricity via electrolysis of water;
- Production from biomass and electricity via hybrid process.

In addition, following synthetic end-product options are evaluated:

- Natural gas (methane);
- Methanol;
- Gasoline.

The combination of these alternatives gives nine basic configurations, each characterised by distinctive plant designs. These configurations are summarised and named in Table 1.

2.1. Thermochemical pathway

Fischer-Tropsch (FT) synthesis is a well known method for producing liquid hydrocarbon fuels from synthesis gas. However, once synthesis gas is produced, other end-product alternatives are also available, including: natural gas (methane), methanol, dimethyl ether (DME) and gasoline [12–15].

Some of the attempts to produce these fuels from biomass-derived synthesis gas have ended in difficulties [16,17], although technical hurdles have since been overcome and synthetic biofuels technology can be currently considered successfully demonstrated at pre-commercial scale [18–20]. Nonetheless, commercial applications are still lacking. The slow commercialisation pace is often attributed to the technology's high specific investment cost, financing gaps on the path from pre-revenue stage to commercial operations, uncertainty about the stability of carbon policies and lack of knowhow in sourcing and processing lignocellulosic biomass.

2.2. Electrochemical pathway

The concept of producing synthetic fuels from carbon dioxide via electrochemical pathway was first proposed in the late 1970s and studied further in the early 80s [21–24]. The early concepts were based on nuclear energy sources and low temperature electrolysis, while more recently the focus has turned to solar and wind using high temperature electrolysis for hydrogen production [25]. The renewed interest in the topic has been fuelled by the improved availability and economics of electricity produced from renewable sources, especially from wind and solar. Synfuels are not currently

Table 1 – Summary of the basic plant configurations considered in this paper. The configurations are identified by a sequence of two letters: first letter identifies the production route and second letter the main product.

	Thermochemical	Hybrid	Electrochemical
Natural gas	TN	HN	EN
Methanol	TM	HM	EM
Gasoline	TG	HG	EG

¹ In this paper, synthetic fuels are defined as fuels manufactured from synthesis gas (CO + H₂) or a mixture of CO₂ + H₂.

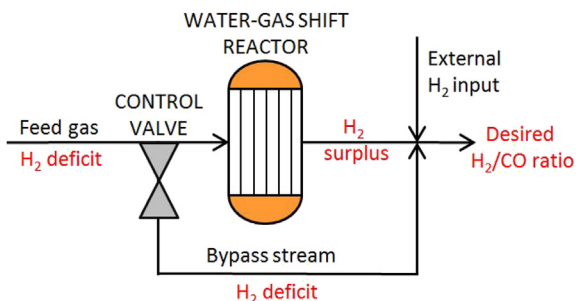
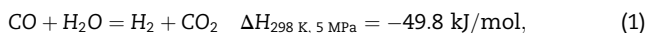


Fig. 1 – Schematic illustration of a configuration suitable for regulating the syngas stoichiometry with a combination of water–gas shift and external hydrogen input.

produced commercially from CO₂ as the main feed, although research is also ongoing to make it possible [26–28].

2.3. Hybrid pathway

Hydrogen and carbon needs to be fed to the synthesis in right proportions to achieve maximal conversion to fuels. Their ratios can be controlled upstream with a reactor that catalyses water-gas shift reaction (1). By controlling the amount of bypass around this reactor, almost any ratio² of H₂/CO can be achieved once the streams are again combined (see Fig. 1).



Another possibility for adjusting the syngas stoichiometry would be to remove the shift reactor completely and directly import the required amount of hydrogen from external sources [29–31]. This approach would also allow more of the syngas' CO to be converted into fuel as losses incurred during the WGS reaction could be avoided. However, such an arrangement requires constant flow of hydrogen leaving little space for flexibility.

The configuration examined in this paper combines the above-described approaches into a one hybrid system that features both a grid connected electrolyser and a WGS step. With such a hybrid approach, a time-variable control over the amount of external hydrogen addition becomes possible. The improved flexibility allows to operate the electrolyser only during times of excess supply of renewable electricity, making it possible to participate in levelling the peaks of time-variable renewable energy production. In principle, no additional technical barriers should be introduced as a result of the hybrid approach, making it possible to deploy such configurations in tandem with synthetic biofuels technology.

3. Technology review and design parameters

This section provides brief descriptions of technologies featured in this paper. The main design parameters are

² The minimum being that of the feed gas' and the maximum that of the shift reactor effluent's.

summarised in Table 2 and discussed in the text below. For a detailed list of modelling parameters and their sources please refer to the appendix. Gasoline is produced from methanol as a separate post-processing step and therefore shares upstream settings with the corresponding methanol plants.

3.1. Biomass to synthesis gas

A simplified block diagram of a plant suitable for the conversion of biomass residues to ultra-clean synthesis gas is shown in Fig. 2. The plant is operated with forest residue chips whose properties are given in Table 3. The wet biomass feedstock is first dried from its initial moisture of 50 wt% to 15 wt% in a belt dryer operated with hot water recovered from the gasification plant. The dried chips are pressurised with lock-hoppers to 0.4 MPa and fed to a circulating fluidised-bed gasification reactor operating at 850 °C. The gasifier is fluidised with equal amounts of steam and oxygen and used to convert wood chips into a raw product gas containing CO, H₂, CO₂, H₂O, CH₄ and small amount of higher hydrocarbons and tars [32].

Before filtration, the gas is cooled³ down to 550 °C to condense alkali metals and to avoid blinding of the filter elements during dust removal [33]. The filtered gas is sent to a catalytic reformer where tars and hydrocarbons are converted to light gases. For plants that produce methanol the tar reformer is designed for maximal methane⁴ conversion (95% at 957 °C), while for plants that produce synthetic natural gas (SNG) the methane conversion is minimised (35% at 850 °C) [33]. The model used to simulate this three-step (gasification, filtration, catalytic reforming) process is validated with experimental data derived from a 0.5 MW_{th} process development unit (PDU) that was run circa 4000 h in pressurised oxygen-blown mode using various wood residues as feedstock [34]. The model itself is described in detail in Refs. [35,36].

After reforming, the gas is cooled down to 260–80 °C, shifted in an adiabatic reactor and cooled further to 200 °C. The syngas is then fed to a two-stage water scrubber that cools the gas down to 60 °C while recovering the heat for feedstock drying and district heating, and then further to 30 °C to condense out the syngas moisture. The dried gas is compressed and fed to an acid gas (CO₂ and sulphur species) separation unit operated with chilled methanol as the washing solvent.

For plants that feature natural gas (SNG) production, syngas is compressed to 1.6 MPa at a one go (inlet pressure to methanation is 1.5 MPa), which allows for 0.1 MPa pressure drop in the AGR. For plants that feature methanol synthesis, syngas is compressed to 8.0 MPa (operating pressure of the methanol converter) in two steps: first to 3.1 MPa before acid gas removal followed by further compression⁵ to 8.0 MPa.

³ In all plants heat is recovered to superheat steam, boil water or generate hot water, depending on the temperature window of cooling.

⁴ This refers to methane that is unavoidably formed during biomass gasification.

⁵ Dividing compression in two parts saves compression work due to the lack of CO₂ in the latter phase.

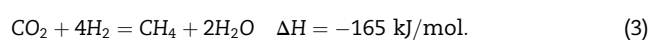
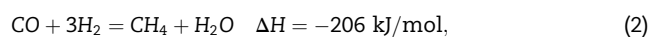
Table 2 – Main design parameters for the examined plant configurations. For sources and additional parameters, see appendix.

Configuration		TN	TM	HN	HM	EN	EM
Band conveyor dryer							
Specific heat consumption	kWh/tH ₂ O _{evap}	1300	1300	1300	1300		
Share of LT heat in belt dryer	%	20	20	20	20		
Moisture in	wt%	50	50	50	50		
Moisture out	wt%	15	15	15	15		
CO₂ capture from fluegas							
CO ₂ purity	mol%					100	100
CO ₂ pressure	MPa					0.1	0.1
Air separation unit (ASU)							
Oxygen purity	mol%	99.5	99.5	99.5	99.5		
Oxygen delivery pressure	MPa	0.105	0.105	0.105	0.105		
Steam/O₂ gasifier							
Pressure	MPa	0.4	0.4	0.4	0.4		
Temperature	°C	850	850	850	850		
Heat loss (HHV)	%	0.8	0.9	0.8	0.9		
Steam/O ₂	kg/kg	1.0	1.0	1.0	1.0		
Carbon conversion	%	98.0	98.0	98.0	98.0		
O ₂ /air/steam mix inlet temp.	°C	203	215	203	215		
CaCO ₃ /Biomass (dry)	wt%	2.0%	2.0%	2.0%	2.0%		
Filter cooler							
Temperature before filtration	°C	550	550	550	550		
Reformer							
Outlet temperature	°C	850	957	850	957		
Heat loss (HHV)	%	1.0	1.0	1.0	1.0		
Steam/O ₂	kg/kg	1.0	1.0	1.0	1.0		
Methane conversion	%	35	95	35	95		
O ₂ /air/steam mix inlet temp.	°C	206	218	206	218		
Sour shift reactor							
Steam/CO at inlet	mol/mol	2.0	2.0				
Reactor inlet temp.	°C	280	258				
Reactor outlet temp.	°C	404	405				
By-pass/syngas feed	mol/mol	0.43	0.68				
H ₂ /CO at exit	mol/mol	3.00	2.03				
Scrubber							
Temperature at inlet	°C	200	200	200	200		
Temp. at stage 1 outlet	°C	60	60	60	60		
Temp. at stage 2 outlet	°C	30	30	30	30		
Syngas compressor							
Syngas pressure at outlet	MPa	1.6	3.1	1.6	3.1		
Acid gas removal							
CO ₂ removal extent	%	98	98	98	98		
Sulphur removal extent	%	99	99	99	99		
Alkaline electrolysis							
Pressure	MPa					0.1	0.1
H ₂ purity	mol%					100	100
O ₂ purity	mol%					100	100
System hydrogen efficiency	%			62	62	62	62
H ₂ /CO after H ₂ addition	mol/mol			3	2.03		
H ₂ /CO ₂ after H ₂ addition	mol/mol					4	3
Syngas conversion							
Inlet pressure to synthesis	MPa	1.5	8.0	1.5	8.0	1.5	8.0
Syngas efficiency	%	>99.5	95	>99.5	95	>99.5	95
Auxiliary boiler							
Boiler fluegas oxygen	mol%	4	4	4	4		
Fluegas to stack	°C	150	150	150	150		

3.2. Synthesis of methane

Methane is synthesised by hydrogenation of carbon oxides over catalysts based on nickel and other metals (Ru, Rh, Pt, Fe and Co) [37], although in practise all commercially available modern catalyst systems are based on nickel due to its favourable combination of selectivity, activity and price [38].

Conversion of synthesis gas to methane can be described with the following reactions:



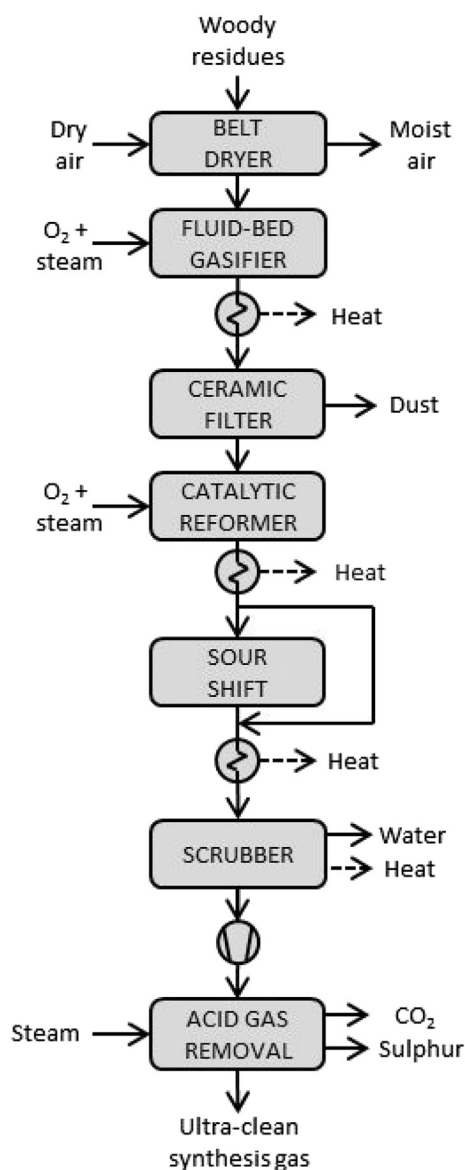


Fig. 2 – Schematic illustration of a thermochemical process capable of converting forest residue chips to ultra-clean synthesis gas.

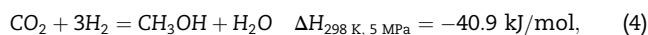
The simulation model developed for this paper is inspired by the high temperature methanation process 'TREMPE', developed and offered by Haldor Topsøe [38,39]. The design features six adiabatic fixed-bed reactors connected in series and equipped with intercoolers. The pressure at the inlet of the first reactor is 1.5 MPa. The inlet syngas is mixed with steam and preheated to 300 °C. The amount of steam addition is chosen to limit temperature increase in the first reactor to 700 °C.⁶ The hot effluent exiting from the first four reactors is cooled to 300 °C before entering to the next reactor in series. Effluent from the fifth reactor is cooled down to condense and separate gas moisture before feeding to the last reactor. Overall conversion of syngas to methane is >99.5% and the

⁶ An alternative design would employ a recycle around the first reactor to limit temperature rise. This design doesn't require steam, but calls for a compressor and electricity to run it.

effluent exits from the system at 1.1 MPa pressure. Equilibrium conversions in the reactors are calculated with Aspen using Soave-Redlich-Kwong (SRK) equation of state model. The recovered heat is used to produce high pressure super-heated steam for the plant's steam cycle.

3.3. Synthesis of methanol

Methanol is synthesised by hydrogenation of carbon oxides over catalysts composed of copper oxide and zinc oxide stabilised with alumina [13]. These catalysts allow the production of methanol at over 99.9% selectivity with higher alcohols as primary byproducts [40]. Synthesis of methanol can be described with the following reactions:



which both are exothermic and result in a decrease in volume [40].

The simulation model developed for this paper is based on the 'low-pressure methanol synthesis', a de facto industrial process for large scale methanol manufacture since the 1960's [41]. In the acid gas removal step, CO₂ is not removed completely, but about 1 mol-% of CO₂ is left to the syngas to improve kinetics of methanol formation. Before treated syngas (make-up gas) can be fed to the methanol process it needs to be compressed further to 8.0 MPa, which is the operating pressure of the methanol converter. The make-up gas is then admixed with the recompressed recycle stream and pre-heated to 260 °C in heat exchange with the hot reactor effluent. The methanol converter design is based on a tubular boiling-water reactor operated at 260 °C and 8.0 MPa. The equilibrium conversions are calculated with Aspen using Soave-Redlich-Kwong (SRK) equation of state model. Any unconverted synthesis gas is separated from the effluent at the reactor exit, recompressed and recycled back to the feed side of the reactor until 95% conversion of CO and H₂ to methanol is achieved. Small amount of unconverted gas needs to be continuously purged to control the build-up of inerts in the methanol loop, which is then transferred to combustion.

The produced raw methanol contains reaction water formed as byproduct of CO₂ conversion. This water (along with small amounts of higher alcohols) can be separated from methanol to achieve a desired product quality. Higher purities can be achieved simply by adding more distillation columns, thus contributing to additional capital and energy costs.

3.4. Synthesis of gasoline

Gasoline is synthesised by a two-step process that involves 1) production of oxygenates from synthesis gas and 2) subsequent conversion of oxygenates to higher hydrocarbons boiling in the gasoline range [42]. These processes may be carried out as separate steps using methanol as the intermediate oxygenate, or in integrated fashion by producing a methanol/DME mixture from syngas that is fed directly to a downstream gasoline converter [43].

Table 3 – Properties of forest residue chips used as feedstock for gasification [75].

Proximate analysis, wt% d.b. ^a	
Fixed carbon	18.3
Volatile matter	80.6
Ash	1.1
Ultimate analysis, wt% d.b.	
Ash	1.1
C	51.48
H	6.0
N	0.2
Cl	0.0
S	0.02
O (difference)	41.2
Other properties	
HHV, MJ/kg	20.67
LHV, MJ/kg	19.34
Bulk density, kg d.b./m ^{3b}	293
Sintering temp. of ash, °C	>1000

^a wt% d.b. = weight percent dry basis.
^b 1 litre batch, not shaken.

The conversion of methanol to gasoline proceeds essentially according to reaction



where $(\text{CH}_2)_n$ represents a wide range of paraffinic and aromatic hydrocarbons produced in the gasoline synthesis step [44]. The process design developed for this paper is based on the conventional two-step methanol-to-gasoline (MTG) process, originally developed by Mobil in the 1970's [45–47]. The conversion of syngas to methanol is analogous to what has been described in section 3.3, although it is assumed that less distillation is required when preparing raw methanol for gasoline production. The MTG process begins by pumping methanol feed to 2.27 MPa followed by vaporisation and superheating to 297 °C in heat exchange with the hot reactor effluent. An adiabatic fixed-bed dehydration reactor is used to convert the feed to an equilibrium mixture of methanol, DME and water. The effluent exits the reactor at 409 °C and 2.17 MPa, is admixed with recycle gas and fed to a second reactor where it is converted to gasoline [47]. A large recycle stream is needed⁷ to limit the outlet temperature of the adiabatic gasoline reactor to 400 °C. To control the build-up of inerts in the synthesis loop some gas needs to be continuously purged from the recycle flow, which is then transferred to combustion.

The equilibrium conversion of methanol to DME and water is simulated with Aspen using Soave-Redlich-Kwong (SRK) equation of state model. Due to the proprietary nature of the process, very little information has been published about the performance of the gasoline reactor, thus complicating the process simulation effort. However, a RYield block was chosen to simulate the gasoline synthesis (with SRK) using product

Table 4 – MTG yield structure for a fixed-bed reactor given per kg of pure methanol input to DME reactor [49].

Component name	Formula	Molar mass	kmol/kgMeOH
Hydrogen	H ₂	2.02	0.00001049
Water	H ₂ O	18.02	0.03137749
Carbon monoxide	CO	28.01	0.00000446
Carbon dioxide	CO ₂	44.01	0.00001390
Methane	CH ₄	16.04	0.00019586
Ethene	C ₂ H ₄	28.05	0.00000473
Ethane	C ₂ H ₆	30.07	0.00005067
Propene	C ₃ H ₆	42.08	0.00002055
Propane	C ₃ H ₈	44.10	0.00042752
1-Butene	C ₄ H ₈	56.11	0.00008593
n-Butane	C ₄ H ₁₀	58.12	0.00019381
i-Butane	C ₄ H ₁₀	58.12	0.00062811
Cyclopentane	C ₅ H ₁₀	70.13	0.00001514
1-Pentene	C ₅ H ₁₀	70.13	0.00014015
N-pentane	C ₅ H ₁₂	72.15	0.00008633
I-pentane	C ₅ H ₁₂	72.15	0.00075797
Gasoline ^a	C ₇ H ₁₆	100.2	0.00283472

^a Gasoline is assumed to be represented as n-heptane (C₇H₁₆).

yield structure (See Table 4) scrutinized by Larson et al. [49] based on the work of Barker et al. [50] and Schreiner [51].

The gasoline reactor effluent is condensed and separated into water, raw gasoline, purge and recycle gas streams. The raw gasoline is then fractionated by distillation to produce finished gasoline blendstock containing less than 2 wt% durene together with a byproduct stream resembling liquefied petroleum gas (LPG) [15,42]. It is assumed that the recovery of waste heat provides the needed utilities for the upgrading, leading to zero net parasitic utilities demand for the area.

3.5. Carbon dioxide capture

Carbon dioxide is available at almost inexhaustible quantities in the atmosphere where it can be captured either directly with an industrial process or indirectly via plant matter [52]. Capturing carbon dioxide from air is fairly easy in chemical sense, but as atmospheric CO₂ is very dilute⁸ (0.04%), the development of a practical system for capturing commercially significant quantities has proved challenging [53].

In a direct air capture (DAC) plant diluted CO₂ is dissolved into a solution or solid sorbent from which a concentrated stream of CO₂ is produced in the regeneration phase. Currently proposed systems are often based on NaOH sorbent followed by regeneration with chemical caustic recovery [4]. The long term cost estimates⁹ for such direct air capture systems are about 115 €/tCO₂ ± 40 €/tCO₂ [52,4]. Despite the high costs, it deserves to be noted that DAC has the unique ability to provide abatement across all economic sectors at a fixed marginal cost [52]. In other words, the cost of DAC represents the upper limit for any conceivable CO₂ abatement strategy.

⁸ 402 parts per million by volume as of May 2014. Based on data collected by continuous atmospheric monitoring at the Mauna Loa Observatory in Hawaii, USA. See www.co2now.org.

⁹ Based on US\$150/tCO₂ ± \$50/tCO₂.

⁷ A molar recycle to fresh feed ratio of 7.5:1 is assumed, a design value for the New Zealand commercial unit [48].

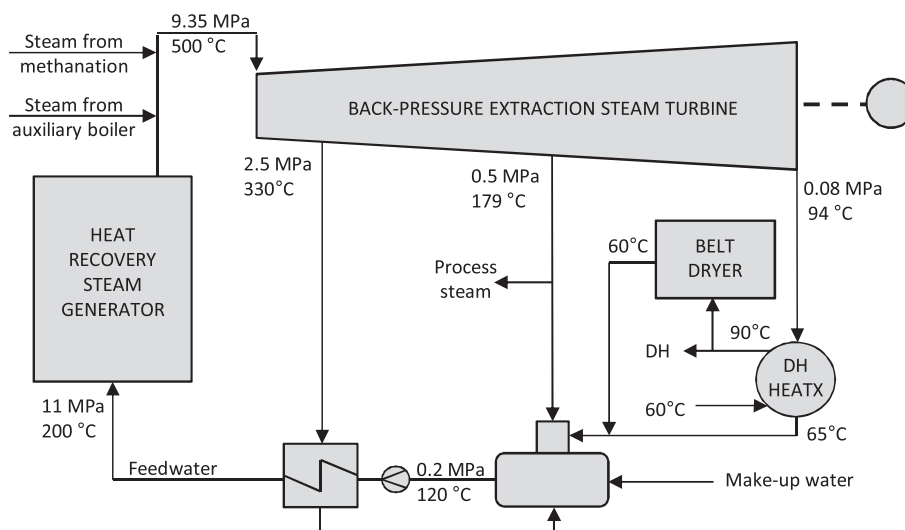


Fig. 3 – Simplified layout of the steam cycle design.

A third possible source for carbon dioxide, in addition to direct and indirect capture from air, is to utilise exhaust CO₂ from industrial plants [24]. Today, carbon dioxide is routinely separated at some large industrial plants and also at several small power plants. The capture costs are estimated to be around 40 €/tCO₂ [54,55] for new supercritical pulverised coal boilers and around 50 €/tCO₂ for new natural gas combined cycle plants [55] both employing an amine-based system for post-combustion CO₂ capture, excluding the cost of transport and storage.¹⁰ However, it is acknowledged that new or improved methods of CO₂ capture have the potential to significantly reduce the cost of capture and required energy use [56].

3.6. Carbon dioxide hydrogenation

Carbon dioxide can be used as a C₁ building block for making organic chemicals, materials and fuels [57]. However, it is considered less favourable feedstock for fuels production than carbon monoxide due to more intensive use of resources (energy, H₂, more reaction steps, etc.) [28]. Presently the use of CO₂ as chemical feedstock is limited to few industrial processes such as urea synthesis and its derivatives, salicylic acid and carbonates [28].

Production of methane from CO₂ via Sabatier reaction (3) is a well known route that can be realised using existing methanation catalysts. In addition, catalysts allowing direct hydrogenation of CO₂ to methanol via reaction (4) have been developed, and pilot-scale plants based on this technology demonstrated [58–62]. However, when methanol is synthesised from a mixture of CO₂ and H₂ instead of syngas, a greatly reduced yield is reported [31,26,63]. In addition, almost one third of the input H₂ is consumed to produce byproduct water.

Although in practise routes via CO are preferred, plant configurations developed for this paper assume that catalyst systems for CO₂ conversion are available and operate close to

equilibrium conversion with the same catalyst productivity than commercial alternatives using carbon monoxide as the main feed.

3.7. Electrolysis of water

Hydrogen can be produced by passing an electric current through two electrodes immersed in water. In the process, water molecules are split to produce oxygen and hydrogen according to the following overall reaction:



Presently the production of hydrogen via electrolysis is mainly limited to small or special applications, while larger quantities are produced by steam reforming of natural gas or other fossil fuels. The most established and commercially available technology is based on alkaline electrolyzers, while proton exchange membrane (PEM) electrolysis and solid oxide electrolysis cells (SOEC) are examples of more advanced and emerging systems [64]. SOEC electrolyzers are most efficient but least developed. PEM electrolyzers are more efficient than alkaline and do not have issues with corrosion or seals as the SOEC systems, but cost more than alkaline systems. Alkaline electrolyzers have the lowest efficiency, but are the most developed and lowest in capital cost [65].

This paper examines hydrogen production via low temperature alkaline water electrolysis.¹¹ The system composes of electrodes, a microporous separator and an aqueous solution of water and 25–30 wt% of potassium hydroxide (KOH) as electrolyte [66]. The liquid electrolyte is not consumed in the reaction, but must be replenished over time to cover losses that occur during hydrogen recovery. Water is decomposed into hydrogen and OH[−] in the cathode. The OH[−] travels through the electrolytic material to the anode where O₂ is formed, while hydrogen is left in the alkaline solution and separated by a gas/liquid separator unit outside the electrolyser cell. Nickel with a catalytic coating, such as platinum, is

¹⁰ Based on US\$50/tCO₂ and US\$60/tCO₂, respectively.

¹¹ Norsk Hydro's Atmospheric Type No. 5040 (5150 Amp DC).

Table 5 – Process simulation results for examined plant configurations.

Configuration		TN	TM	HN	HM	EN	EM
Carbonaceous inputs							
Biomass to dryer (moist. 50 wt-%)	MW (LHV)	100	100	100	100		
Biomass to gasifier (moist. 15 wt-%)	MW (LHV)	112	112	112	112		
Biomass dry feed	kg/s	5.9	5.9	5.9	5.9		
Captured and concentrated CO ₂	kg/s					3.6	4.36
Oxygen balance							
On-site consumption, t/d		280	324	280	324		
Gasifier oxygen input	kg/s	2.3	2.3	2.3	2.3		
Reformer oxygen input	kg/s	0.9	1.4	0.9	1.4		
Air separation unit output, t/d		280	324	280	324		
Electricity balance							
On-site consumption, MW		−9.1	−12.4	−10.3	−13.9	−4.8	−7.3
Oxygen production	MW	−3.1	−3.6	−3.1	−3.6		
Oxygen compression	MW	−0.6	−0.7	−0.6	−0.7		
Feed screw and lock-hopper pres.	MW	−0.2	−0.2	−0.2	−0.2		
Feed drying	MW	−0.7	−0.7	−0.7	−0.7		
Syngas compression	MW	−3.0	−5.2	−2.5	−5.1		
Acid gas removal	MW	−1.0	−0.9	−0.8	−0.8		
Electrolytic H ₂ compression	MW			−1.9	−1.6	−3.7	−5.4
CO ₂ compression	MW					−0.9	−1.6
Synthesis	MW	0.0	−0.3	0.0	−0.4		
Power Island (all blowers + pumps)	MW	−0.2	−0.3	−0.2	−0.3		
Miscellaneous	MW	−0.4	−0.6	−0.4	−0.6	−0.2	−0.3
Turbine gross output, MW		7.7	8.3	8.4	8.6		
Steam balance							
On-site consumption (excl. synthesis), kg/s		7.2	8.0	6.7	7.6		
Gasifier	kg/s	2.4	2.4	2.4	2.4		
Reformer	kg/s	0.9	1.4	0.9	1.4		
AGR solvent regeneration	kg/s	1.2	1.1	0.9	0.9		
Deaerator	kg/s	1.2	1.4	1.1	1.3		
Economiser	kg/s	1.5	1.7	1.4	1.7		
Gross production, kg/s		12.5	16.1	13.3	17.0		
Gasification plant (9.35 MPa, 500 °C)	kg/s	9.3	10.6	8.6	10.2		
Auxiliary boiler (9.35 MPa, 500 °C)	kg/s	0.6	2.0	0.6	2.4		
Admission steam (4.3 MPa, 255 °C)	kg/s		3.5		4.4		2.2
Admission steam (9.35 MPa, 500 °C)	kg/s	2.7		4.1		2.9	
Turbine extractions, kg/s		7.2	4.5	6.7	3.2		
HP steam (2.5 MPa, 330 °C)	kg/s	1.5	1.7	1.4	1.7		
IP steam (0.5 MPa, 179 °C)	kg/s	5.7	2.8	5.3	1.5		
Turbine back pressure	MPa	0.08	0.08	0.08	0.08		
By-products							
Char							
Heating value (LHV)	MJ/kg	7.0	7.0	7.0	7.0		
Energy	MW (LHV)	2.1	2.1	2.1	2.1		
Purge gas							
Heating value (LHV)	MJ/kg		11.5		13.0		14.5
Energy	MW (LHV)		4.9		6.1		3.7
Alkaline electrolysis							
Electricity input	MW			65.8	35.4	129.6	116.1
Hydrogen output	kg/s			0.34	0.18	0.67	0.60
Oxygen output	kg/s			2.70	1.45	5.32	4.76
Energy outputs							
Methanol	MW (LHV)		60.0		78.3		60
SNG (methane)	MW (LHV)	66.7		100.3		66.7	
Net electricity output	MW _e	−1.4	−4.0	−67.7	−40.7	−134.4	−123.4
District heat (from steam cycle)	MW _{th}	3.1	12.3	8.0	16.6		
District heat (from methanation)	MW _{th}	7.4		13.3		12.1	

the most common cathode material, while for the anode nickel or copper metals coated with metal oxides, such as manganese, tungsten or ruthenium, are used [65].

Commercial systems are typically run with current densities in the range of 100–300 mA/cm². The product hydrogen

and oxygen can be assumed to be of 100% purity due to the very low concentration of contaminants [67]. The system efficiency of an alkaline electrolyser, defined as hydrogen output (LHV) divided by electrical energy consumed by the electrolysis system, is set to 62% (74% HHV) [67].

3.8. Steam system

All thermochemical and hybrid plant configurations considered in this paper feature a back-pressure steam turbine design that co-generates electricity and district heat (DH) and has live steam parameters of 9.35 MPa and 500 °C. A simplified layout of the steam cycle is illustrated in Fig. 3. Steam is produced from syngas cooling and in the auxiliary boiler where unconverted carbon from gasifier (filter dust) and unconverted syngas from synthesis (if available) are combusted. Plant configurations that feature syngas methanation produce also superheated steam as a byproduct that is directly usable in the turbine as admission steam. The synthesis exotherm from the methanol plants is utilised to produce saturated admission steam at 4.3 MPa and 255 °C that is directly used to satisfy part of the plant's on-site steam consumption. The first steam extraction point from the turbine is fixed at 2.5 MPa and 330 °C and used to preheat the HRSG feedwater to 200 °C. The second extraction point is fixed at 0.5 MPa¹² and 179 °C and used to supply steam for the gasifier, reformer, deaerator and AGR solvent regeneration. Rest of the steam is extracted at the turbine's back-pressure (0.08 MPa), condensed and used to produce hot water at 90 °C. The hot water provides heat for drying the wet biomass feed and the rest is sold to a near-by district heating grid.

All the examined plant configurations are designed as self-sufficient in terms of heat and steam, while electricity is balanced with the electric grid. Design choices have notable impact to the electricity and steam outputs of the steam system. In this paper, the goal for the steam system design has been simplicity and low capital cost, even if at the cost of lower performance. Comparable results for synthetic biofuel plants equipped with a higher performance (and supposedly more expensive) steam system are available in Ref. [36].

4. Performance analysis

4.1. Mass and energy balances

Mass and energy balances have been simulated for all examined plant configurations. The thermochemical and hybrid plants are based on 100 MW (LHV) input of wet biomass, while fuel output for the electrochemical plants is set equal with the corresponding thermochemical plants.

The main simulation results are summarised in Table 5. With thermochemical plant configurations, 66.7 MW of natural gas or 60.0 MW of methanol can be produced from 100 MW of wet biomass. Producing the same fuel outputs with electrochemical configurations requires 129.6 MW of electricity and 3.6 kg/s of CO₂ (natural gas plants), or 116.1 MW of electricity and 4.36 kg/s of CO₂ (methanol plants). When syngas production is boosted with the maximum amount¹³ of electrolytic hydrogen (hybrid configurations) natural gas output increases by 50 % to 100.3 MW and methanol output by 31 % to 78.3 MW. The greater increase in natural gas

production is due to the larger stoichiometric hydrogen requirement in relation to methanol production.

The net electricity output is negative for all examined plant configurations. However, the on-site electricity consumption of thermochemical plants is fairly comparable with on-site generation and net electricity surplus could be achieved with a (more expensive) steam cycle designed for higher performance as discussed in section 3.8. For the electrochemical and hybrid plants, electricity consumption of the alkaline electrolysis clearly dominates electricity balance, leading to deeply negative net outputs. As already noticed, more electricity (i.e. hydrogen) is required to produce natural gas than methanol. However, the difference in net electricity requirement between methanol and natural gas production is smaller for the electrochemical than for hybrid configurations. This can be explained by the increased role of compression work in electrochemical plants (feed gases starting at atmospheric pressure, while gasifier operates at 0.4 MPa) that level down differences in electricity consumption during syngas conversion.¹⁴

In addition to synfuel, most plant designs co-produce district heat at 90 °C. The district heat outputs for methanol plants are 12.3 MW (TM) or 16.6 MW (HM) depending on the configuration. For natural gas configurations some DH can also be produced from the methanation area in addition to steam cycle. The combined DH output from such plants is 10.5 (TN), 21.3 (HN) or 12.1 MW (EN) depending on the configuration. As the electrochemical configurations omit steam cycle, district heat is available only from plants producing natural gas.

Differences in steam system designs are clearly visible from the simulation results: Less intermediate pressure extraction steam from turbine is required for methanol plants as part of the on-site consumption is satisfied directly with saturated steam produced from the methanol reaction's exotherm. In addition, auxiliary boiler's steam output is smaller for natural gas plants than for methanol plants because purge gas is not formed in methanation.

The gasifier's oxygen consumption is constant for all configurations, but the amount of oxygen required in reforming depends on the targeted methane conversion. For purely thermochemical plants, the combined oxygen requirement is 3.2 kg/s (natural gas) or 3.7 kg/s (methanol). Oxygen is also produced as a co-product with hydrogen in configurations that feature alkaline electrolysis. The net oxygen output for hybrid configurations is –2.3 kg/s (HM) or –0.5 kg/s (HN) and for the electrochemical plants, where oxygen is not consumed by the process, the net oxygen output is 4.8 kg/s (EM) and 5.3 kg/s (EN).

In this paper, gasoline production is treated as a post-processing step that may or may not take place at the same site with methanol production. Simulation results for the methanol-to-gasoline (MTG) conversion step are given in Table 6 together with the overall performance results for the total conversion path from feedstock to gasoline via methanol. Thermochemical and electrochemical configurations both produce 60.0 MW of methanol from which 51.8 MW of

¹² 0.1 MPa higher than gasification pressure to allow pressure drop for the inlet valves.

¹³ Complete bypass of the water-gas shift reactor.

¹⁴ Methanol production requires less hydrogen than methanation but takes place at much higher pressure.

gasoline and 6.1 MW of LPG can be further produced via MTG process. From the 78.3 MW of methanol produced by the hybrid process 67.6 MW of gasoline and 7.9 MW of LPG can be produced. Some high pressure saturated steam is also generated from the gasoline reaction's exotherm, which can be sold as process steam or utilised as an admission steam if a steam cycle is available nearby.

4.2. Energy input breakdowns

Fig. 4 illustrates energy input breakdowns for each of the examined plant configurations. The simulation results, presented in Tables 5 and 6, have now been rescaled to fit a situation where the fuel output is 200 MW (LHV) for all plants. For the thermochemical configurations, following amounts of wet biomass is required to produce 200 MW of synthetic fuel: 300 MW (TN), 333 MW (TM) or 386 MW (TG). A notable drop in biomass requirement is achieved with hybrid configurations where biomass is partly replaced with electricity. The feedstock requirements are: 199 and 131 MW (biomass and electricity) for natural gas, 255 and 90 MW for methanol, 296 and 105 MW for gasoline production. For pure electrochemical designs, where biomass is fully replaced with electricity and carbon dioxide, 200 MW of synthetic fuel can be produced from 389 MW (EN), 387 MW (EM) or 448 MW (EG) of electricity and 10.8 kg/s (EN), 14.5 kg/s (EM) or 16.8 kg/s (EG) of carbon dioxide.

5. Cost analysis

The capital cost and the cost of producing fuel are estimated for each of the modelled plant configurations using a set of

simplified economic assumptions. The overall economics are evaluated under alternative feedstock price assumptions from the perspective of a synthetic fuel producer in terms of euros (€) per gigajoule (GJ). The value of the analysis lies not in the absolute accuracy of individual results, but in the fact that all plant designs have been consistently evaluated under the same set of technical and economic assumptions.

It is acknowledged that the use of 'unproven technologies' in a plant is likely to cause increased capital costs and decreased plant performance. In fact, conventional estimating techniques, like the one used here, have been found out to routinely understate the costs of innovative technologies [68]. Thus, it is highly likely for the first commercial scale installations of these plants to be more expensive than estimated here, although the probable level of misestimation is difficult to assess in advance. In any case, the aim of this paper is to evaluate and understand the long-term commercial viability of the examined plant designs, i.e. when all plant components have already reached commercial maturity. Methods suitable for the analysis of first-of-a-kind plant costs have been proposed, perhaps the most famous being that based on empirical formulae developed by RAND Corporation [68], but carrying out such analysis is out of the scope of this paper.

5.1. Scale of operations

The overall costs of synthetic fuel production are subject to economies of scale, which creates an incentive to build large conversion plants. However, due to limitations in the availability of biomass feedstock, biofuel plants are confined to a much smaller scale than modern synfuel plants based on coal, shale or natural gas conversion. For example, the largest pulp and paper mills in Europe process annually about one million tons of dry biomass that relates to about 600 MW of constant energy flow,¹⁵ which in this paper is considered as the maximum size of biomass conversion plants.

Another possible way of estimating proper scale for a biomass conversion plant would be to consider maximal byproduct utilisation. In northern Europe, a typical annual heat demand for district heating networks, situated at or close to wooded territories, range from 450 to 1700 GWh/a with peak loads between 150 and 650 MW.¹⁶ However, a better indicator for scale would be the minimum continuous load (summer load), which ranges from 50 to 150 MW [69].

In observance of these limitations, fuel output for all examined plants is set to 200 MW (see Table 4), which is large enough to attain some economies of scale, while keeping feedstock requirements under practical limits and ensuring complete utilisation of byproduct heat.¹⁷

Table 6 – Simulation results for upgrading the methanol to synthetic gasoline with the MTG process.

Configuration		TG	HG	EG
Results for gasoline synthesis				
Methanol input	MW	60.0	78.3	60.0
Inlet pressure to synthesis	MPa	2.3	2.3	2.3
Outlet pressure from synthesis	MPa	1.7	1.7	1.7
DME reactor inlet temp.	°C	297	297	297
DME reactor outlet temp.	°C	407	407	407
Once-through MeOH conversion	%	82	82	82
MTG reactor outlet temp.	°C	400	400	400
Recycle/MeOH	mol/mol, wet	7.5	7.5	7.5
Purge gas energy flow	MW	3.0	3.9	3.0
Total MeOH conversion	%	100	100	100
Gasoline LHV	MJ/kg	44.7	44.7	44.7
LPG LHV	MJ/kg	45.9	45.9	45.9
Net electricity output	MW	−0.2	−0.2	−0.2
Net steam output	kg/s	2.2	2.8	2.2
Results for gasoline plant configuration				
Gasoline energy	MW	51.8	67.6	51.8
LPG energy	MW	6.1	7.9	6.1
Net electricity output	MW	−4.2	−40.9	−123.6
District heat (from steam cycle)	MW	12.3	16.6	
Net steam output	MW	2.2	2.8	2.2

¹⁵ Assuming 8000 annual operating hours and 8.6 MJ/kg lower heating value for forest residues at 50 wt% moisture.

¹⁶ The data is based on municipal DH networks situated in eastern Finland sampled from Ref. [69].

¹⁷ When fuel output is set to 200 MW, biomass feedstock requirements for the examined plant configurations vary from 199 to 386 MW (TG having the largest) and DH outputs from 0 to 43 MW (HN having the largest).

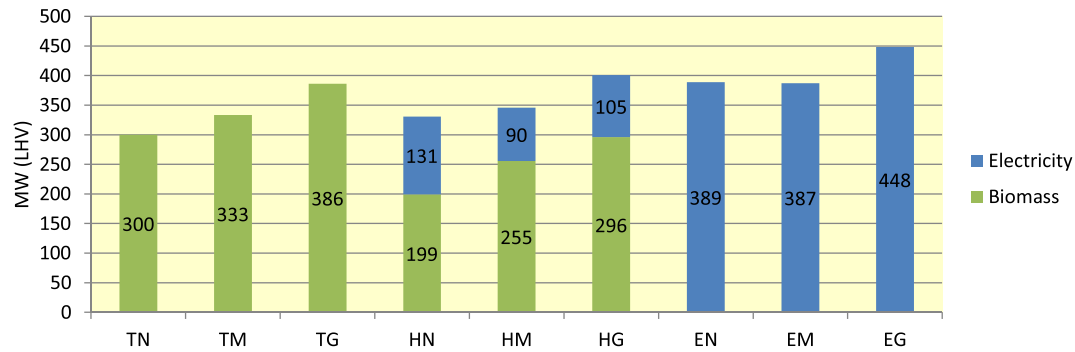


Fig. 4 – Feedstock requirements for all examined plant configurations producing 200 MW (LHV) of fuel. Only electricity used in the electrolyser is included.

5.2. Reference equipment cost database

Capital cost estimates provide the basis for evaluating prospective economics of synthetic fuel production. The estimates are based on a self-consistent set of component level capital cost data assembled using literature sources, vendor quotes and discussions with industry experts. When data for a given equipment has been unattainable, costs have been estimated based on similar equipment and engineering judgement. All reference costs in the database (see Table 9) have been escalated to correspond with 2010 euros using Chemical Engineering's Plant Cost index¹⁸ (CEPCI) to account for inflation.

Total plant cost (TPC) is defined as the "overnight" capital investment required to construct a plant, and total capital investment (TCI) is TPC plus interest during construction. The TPC includes all main equipment (with initial catalyst loadings) plus installation (labour), indirect costs (engineering and fees), project contingency and unscheduled equipment. These cost items are estimated with cost factors that are reported as a fraction of the (installed) equipment costs. Two different cost factor sets have been employed: higher for bare equipment and lower for whole subprocesses delivered on turn-key basis. Costs for unscheduled equipment are added to represent pumps, blowers and other small equipment not included in the cost database. Following values for cost factors¹⁹ are used in this paper:

- Installation costs: low 15%, high 50%;
- Indirect costs: low 10%, high 15%;
- Project contingency: low 10%, high 30%;
- Unscheduled equipment: 10%.²⁰

The allocation of the above factors with reference equipment costs is shown in Table 9.

Table 7 – Financial parameters employed in the cost analysis.

Financial parameters	
Annuity factor (10%, 20 a)	0.12
Annual O&M cost factor	0.04 ^a
Annual operating hours	8000
Interest during construction	5% ^a
Investment support, M€	0
Values of inputs/outputs	
Biomass residue chips, €/GJ	5
District heat, €/GJ	8
Fuel gas, €/GJ	10
LPG, €/GJ	12
Electricity, €/GJ	14
Water, €/t	0
Oxygen, €/t	27
Steam, €/t	30
Carbon dioxide, €/t	40

^a Fraction of Total Capital Investment.

5.3. Methodology and parameters

Accurate information of absolute equipment costs are often unavailable in the open literature. In observance of this limitation, the aim is set rather to estimate how relative costs compare among alternative systems with a reasonably high degree of confidence. The analysis lends from methodologies put forth and discussed in Refs. [72,6,70].

The annual capital charges are calculated from the TCI using 0.12 annuity factor, which corresponds with 10% interest and 20 years lifetime. The yearly operating and maintenance (O&M) costs²¹ are assumed to be 4% of the TCI [73] and the plants are expected to run 8000 h per year (91% on-stream factor). Biomass is valued at 5 €/GJ (18 €/MWh).²² Costs

¹⁸ For more information, see: www.che.com/pci.

¹⁹ Cost factors, except unscheduled equipment, are based on averaged values for selected equipment derived from "Exhibit 3-41" Case 2 of Ref. [70].

²⁰ Factor taken from Wan [71].

²¹ Following breakdown is assumed for the O&M: Personnel costs 0.5%, Maintenance & insurances 2.5%, Catalysts & chemicals 1%.

²² Price for forest residues (thinnings, stumps and slash) and industry byproduct (sawdust and bark) traded for energy or heat production purposes in Finland. Source: FOEX Indexes Ltd.

Table 8 – Breakdown of LCOFs under economic assumptions summarised in Table 7.

Configuration	TN	HN	EN	TM	HM	EM	TG	HG	EG	
Biomass	7.5	5.0		8.3	6.4		9.7	7.4		€/GJ
CO ₂			2.2			2.9			3.4	€/GJ
Electricity	0.3	9.4	28.2	0.9	7.3	28.8	1.1	8.5	33.4	€/GJ
District heat	–1.3	–1.7	–1.5	–1.6	–1.7		–1.9	–2.0		€/GJ
Steam							–1.3	–1.2	–1.3	€/GJ
Oxygen		–0.7	–2.2		–0.5	–2.1				€/GJ
Fuel gas							–0.6	–0.6	–0.6	€/GJ
LPG							–1.4	–1.4	–1.4	€/GJ
O&M	2.8	2.7	2.5	3.2	3.2	2.6	4.2	4.2	3.6	€/GJ
Capital charges	8.3	8.0	7.6	9.7	9.7	7.9	12.7	12.7	10.9	€/GJ
Levelised production cost	17.7	22.7	36.8	20.6	24.4	40.0	22.6	27.6	48.0	€/GJ
	63.6	81.7	132.6	74.1	87.7	144.1	81.3	99.4	173.0	€/MWh

related to buying or selling other feedstocks, byproducts²³ and utilities are summarised with other financial assumptions in Table 7.

The maximum size S_{max} of a single gasification train, containing all equipment between dryer and acid gas removal unit, is set to 200 MW of biomass feedstock (as received, LHV). For plant configurations having size $S > S_{max}$, two gasification trains are installed in parallel, both the size of $S/2$. Since multiple trains typically share some auxiliary equipment and installation labour for two equal units is likely to be less than twice the cost for a single unit, it is assumed that the installed cost for additional train is somewhat less than for the first train [73]. This idea is captured in the analysis with the following equation:

$$C_{mult} = C \times n^{0.9}, \quad (8)$$

where C_{mult} is the joint cost of multiple trains, C the cost of first train and n the number of trains [73].

5.4. Capital cost estimates

Individual cost scaling exponents (k) have been used to scale the reference capital costs (C_0) to a capacity that corresponds with simulation results (S) using a following relation:

$$C = C_0 \times \left(\frac{S}{S_0}\right)^k, \quad (9)$$

where S_0 is the scale of reference equipment and C the cost of equipment at the size suggested by simulation. The aggregated Total Capital Investment estimates, based on an underlying component-level costing, are shown in Table 10. The results are based on a “feasibility study” level of design engineering, which usually carries an accuracy of $-15\%/+30\%$ [70].

The TCIs range from 363 to 611 M€ among the cases analysed. The gasoline configurations (TG, HG and EG) are the most capital intensive as they include all the components of a methanol plant²⁴ plus equipment required for the conversion of methanol to gasoline. For all end-products,

²³ The value of byproduct oxygen from the electrolyser is based on a levelised cost of oxygen calculated for a cryogenic air separation unit featured in plant configuration TM.

²⁴ With the exception of methanol distillation, that is cheaper for gasoline configurations, because water does not have to be completely removed from the MTG unit's feed.

thermochemical configurations have the highest and electrochemical the lowest TCIs. The TCIs for hybrid plants are only slightly lower than those for corresponding thermochemical plants. Differences in TCIs are smaller among plants producing natural gas (TN, HN and EN) than other fuels. This can be explained by the higher hydrogen requirement in comparison to methanol production (3 instead of 2 in CO hydrogenation and 4 instead of 3 in CO₂ hydrogenation) that increases the size and cost of alkaline electrolysis and H₂ compression systems in natural gas configurations.

5.5. Production cost estimates

The levelised cost of fuel (LCOF) production is then evaluated according to the following equation:

$$LCOF(\text{€/GJ}) = \frac{F + E + C + O - R}{P}, \quad (10)$$

where

- F is the cost of feedstock (biomass residues and carbon dioxide),
- E the cost of electricity,
- C the capital charges,
- O the operating and maintenance costs and
- R the revenue from selling byproducts (district heat, steam, electrolytic oxygen, purge gas and LPG).²⁵

The sum of these annual costs (€/a) is divided by P , which is the annual output of fuel (GJ/a) from the plants. When defined in this way, the LCOF (€/t) indicates the break-even price for the produced fuel under the technical and economic parameters assumed.

The contribution of different cost categories to the total LCOFs are shown in Table 8. Among the cases analysed, the LCOFs range from 18 to 48 €/GJ (64–173 €/MWh). For thermochemical configurations (TM, TN, TG) the capital charges and cost of biomass feedstock make about an equal contribution to the LCOF, whereas for hybrid plants (HM, HN, HG) the main contributions come from capital charges, biomass feedstock and electricity. Electricity clearly dominates the production costs in the electrochemical cases and revenue

²⁵ Steam, purge gas and LPG are sold only from MTG plant, oxygen is sold only from plants that feature electrolysis.

Table 9 – Reference equipment costs and cost factors employed in estimating Total Plant Costs.

Cost component	Cost scaling parameter	S ₀	UEC, M€	IC	C ₀ , M€	IDC	PC	k	Notes
Civil works (buildings and structures)	Feedstock, MW _{th} (LHV, AR)	300			12.8	10%	30%	0.85	a
ASU (stand alone) incl. compressor	Oxygen output, t/h	76.6			36.8	10%	10%	0.50	b
Feedstock handling	Feedstock, MW _{th} (LHV, AR)	157			5.3	10%	10%	0.31	c
Belt dryer	Water removal, kg/s	0.342			1.9	10%	10%	0.28	d
Pressurised fluidised-bed gasifier	Dry matter input, kg/s	17.8	25.1	50%	37.7	15%	30%	0.75	a
Ceramic hot-gas filter	Syngas input, kmol/s	1.466	5.9	15%	6.8	15%	30%	0.67	a
Catalytic POX reformer	Syngas input, kmol/s	2.037	14.5	50%	21.8	15%	30%	0.67	a
WGS reactor stage	Feed to gasifier, MW _{th} (LHV)	1377			12.6	15%	30%	0.67	e
Scrubber	Syngas input, kmol/s	1.446			5.2	15%	30%	0.67	a
Syngas compressor	Compressor work, MW _e	10			5.0	15%	30%	0.67	f
CO ₂ compressor	Compressor work, MW _e	10			5.0	15%	30%	0.67	f
O ₂ compressor	Compressor work, MW _e	10			5.7	15%	30%	0.67	f
H ₂ compressor	Compressor work, MW _e	10			5.7	15%	30%	0.67	g
Rectisol incidentals compression	Compressor work, MW _e	10			5.0	15%	30%	0.67	f
Rectisol, sep. capture of CO ₂ and H ₂ S	Nm ₃ /hr (NTP) input sourgas	200000	49.3	15%	56.7	15%	30%	0.63	h
Alkaline electrolysis	Electricity input, MW _e	223.5			121.9	15%	10%	0.93	i
Heat recovery steam generation system	Heat transferred, MW _{th}	43.6	5.2	15%	6.0	15%	30%	0.80	b
Auxiliary boiler & fluegas treatment	Boiler input, MW _{th}	5.9	5.1	15%	5.9	10%	10%	0.65	j,k
Steam turbine unit	Power out, MW _e	15.2	6.8	15%	7.8	10%	10%	0.85	j,l
CHP equipment	Power out, MW _e	15.2	4.1	15%	4.7	10%	10%	0.85	j,m
Other steam cycle equipment	Power out, MW _e	15.2	6.3	15%	7.3	10%	10%	0.85	j,n
Guard beds	Syngas, MW _{th}	260	5.2	15%	6.0	10%	10%	0.85	o
Methanol loop	Methanol, MW (LHV)	210	28.3	15%	32.5	10%	10%	0.67	o
Methanol distillation (minimal)	Methanol, MW (LHV)	210	4.2	15%	4.8	10%	10%	0.88	o, p
Methanol distillation (chem-grade)	Methanol, MW (LHV)	210	12.6	15%	14.5	10%	10%	0.88	o, p
Methanation	Methane, MW (LHV)	210	28.3	15%	32.5	15%	30%	0.67	q
MTG DME reactor	Gasoline, bbl/day	16 667			45.3	15%	30%	0.67	r
MTG gasoline reactor	Gasoline, bbl/day	16 667			101.2	15%	30%	0.67	r
MTG gasoline finisher	Gasoline, bbl/day	5556			8.2	15%	30%	0.67	r

Note: C₀ is the cost of a installed reference equipment of size S₀ in 2010 euros and k is the cost scaling factor. UEC stands for uninstalled equipment cost, IC for installation costs, IDC for indirect costs and PC for project contingency.

a – Author's estimate.

b – Taken from Larson et al. [76].

c – Costs taken from Ref. [77]. Scaling exponent calculated from two different size handling systems using feedstock energy flow as scaling parameter.

d – Reference capacity and costs taken from Ref. [77]. Scaling exponent calculated based on information on two different size dryers using water removal rate as scaling parameter. Drying capacity is increased by extending the dryer, which results in unusually low scaling factor (middle parts are fairly affordable in comparison to the ends of the dryer).

e – Extracted from Kreutz et al. [72]. This cost is for two-stage equipment that includes balance of plant (15%) and indirect costs (15%). It is assumed that a single-stage adiabatic sour shift reactor is 40% of the cost of a two-stage system (see Ref. [73]). Balance of plant and indirect costs have been removed.

f – Taken from Kreutz et al. [72].

g – It is likely that H₂ compressor is more expensive than O₂ compressor of similar size (electricity usage), but in the lack of reliable cost data an equal cost is assumed.

h – This cost is for a Rectisol system that separates CO₂ and H₂S to separate streams (separate column for each compound). Taken from Liu et al. [73].

i – Cost is for an alkaline electrolysis installation containing 96 individual NorskHydro's No. 5040 atmospheric electrolyzers each having a capacity of 2.3 MW. Cost taken and scaling exponent fitted with data from Floch et al. [78].

j – Costs based on Thermoflow PEACE equipment cost estimator and discussions with experts at ÅF-Consult.

k – Includes boiler and related systems such as air preheaters, fans, ducts, stack, fabric filter et cetera.

l – Includes turbine, generator and electrification related to the delivery.

m – Includes items such as water cooled condenser, district heaters, deaerator et cetera.

n – Includes items such as tanks, pumps, fans, makeup water system, fuel & ash handling systems et cetera.

o – Taken from Refs. [79], originally based on a quotation from Haldor Topsøe in September 2003. Recalculated.

p – Cost (down) scaling factor from Wan [71].

q – Methanation system is assumed to have same cost as methanol loop (i.e. distillation equipment excluded) with equal fuel output.

r – Taken from Larson et al. [49]. Approximately one third of the raw gasoline from MTG reactors is processed through finisher.

received from selling byproducts is small in comparison to the main cost items for all cases analysed. For each product, thermochemical plants have the lowest and electrochemical plants the highest LCOFs with hybrid configurations placing in

between the two. For a given route, natural gas (SNG) is the cheapest and gasoline the most expensive to produce. It is interesting to note that for a given product, the configuration that has the highest investment has the lowest production

Table 10 – Total capital investment estimates for the examined plant configurations each producing 200 MW (LHV) of fuel.

Configuration	TN	HN	EN	TM	HM	EM	TG	HG	EG
Installed equipment cost									
Civil works	13	9		14	11		16	13	
Oxygen production	25	20		28	25		30	27	
Feedstock pretreatment	15	11		16	14		16	15	
Gasification	42	28		45	37		51	41	
Hot-gas cleaning	33	21		37	31		41	34	
CO shift	6	4		6	5		7	6	
Scrubber	7	5		8	7		9	8	
Syngas compression	7	4		10	8		11	9	
Acid gas removal	39	28		42	34		46	37	
Alkaline electrolysis		74	204		53	203		60	233
HRS, boiler and steam cycle	45	33		54	44		60	49	
Additional syngas compression				3	3		3	3	
H ₂ compression		3	6		3	8		3	9
CO ₂ compression			2			3			4
Guard beds and methanol loop				37	37	37	41	41	41
Distillation (minimal)							5	5	5
Distillation (chemical-grade)				14	14	14			
Guard beds and methanation	37	37	37						
MTG synthesis							67	67	67
Sum of installed equipment cost	269	278	249	314	326	266	403	418	359
Indirect costs	34	25	37	40	34	37	53	46	52
Contingency	52	36	34	60	50	29	85	74	52
Unscheduled equipment	27	28	25	31	33	27	40	42	36
Total plant cost	381	367	345	445	442	359	581	581	498
Interest during construction	19	18	17	22	22	18	29	29	25
Total capital investment	400	385	363	467	464	377	611	610	523

cost and vice versa. This can be explained by the relative affordability of biomass residues in comparison to electricity under the assumptions made in this paper. The main results have been visualised in Fig. 5 that summarises TCIs and LCOFs for each of the examined plant configurations.

5.6. Sensitivity analysis

Cost implications of alternative feedstock prices are then investigated. Fig. 6 shows production costs for all examined plant configurations as a function of electricity price while keeping the cost of biomass and carbon dioxide constant at 5 €/GJ (18 €/MWh) and 40 €/t, respectively. All gasoline (MTG) plants are indicated with blue (in web version), methanol plants with red (in web version) and natural gas (SNG) plants

with green (in web version) lines. In addition, the lines are continuous for thermochemical plants, dashed for hybrid plants and dotted for electrochemical plants. As expected, the LCOFs for the thermochemical plants are fairly insensitive to changes in the cost of electricity due to their low net electricity consumption. When the price of electricity changes by 1 €/GJ, it causes a change in the LCOF that is, on average, 0.6 €/GJ for hybrid and 2.2 €/GJ for electrochemical plants. In addition, it can be seen that the costs for hybrid plants are lower in comparison to corresponding thermochemical plants producing the same fuel when the price of electricity is below 6 €/GJ (22 €/MWh). For electrochemical configurations this price threshold is 4 €/GJ (14 €/MWh). It should be noted that these required threshold values are markedly lower than the

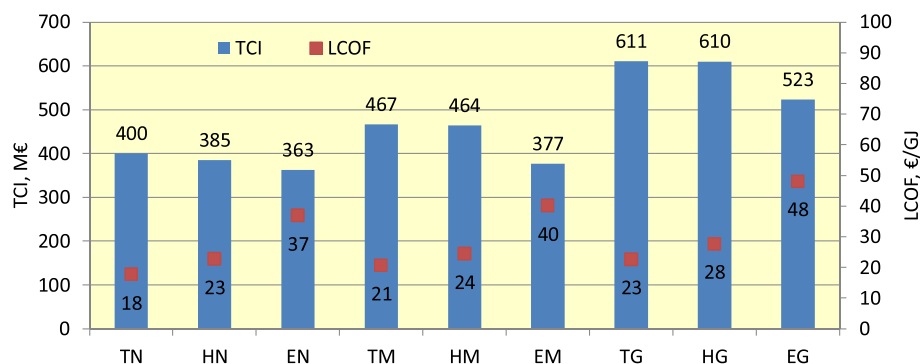


Fig. 5 – Summary of total capital investments (TCIs) and fuel production costs (LCOFs) for plants producing 200 MW (LHV) of fuel.

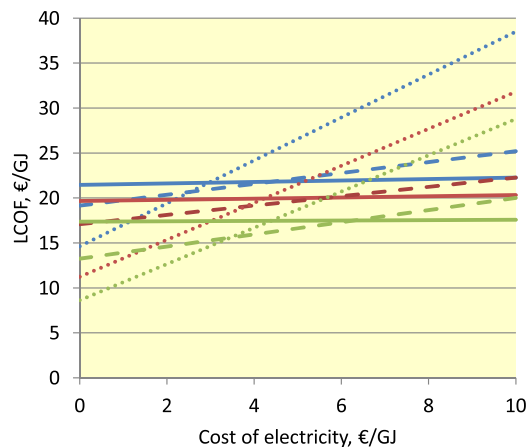


Fig. 6 – Levelised cost of fuel production (LCOF) for the examined plant configurations as a function of the electricity price.

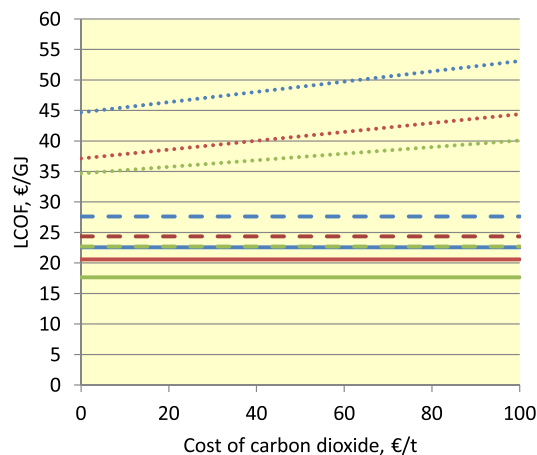


Fig. 7 – Levelised cost of fuel production (LCOF) for the examined plant configurations as a function of the carbon dioxide price.

current EU27 average prices 16–20 €/GJ (58–72 €/MWh) paid by the chemical industry.²⁶

A similar analysis is performed as a function of carbon dioxide price while keeping the cost of biomass and electricity constant at 5 and 14 €/GJ, respectively (see Fig. 7). The costs of thermochemical and hybrid plants remain unchanged (because external CO₂ not used), but every 10 €/t change in the price of CO₂ causes, on average, a 0.7 €/GJ change in the LCOF for electrochemical plants. Somewhat surprisingly, even zero cost carbon dioxide would not be enough to make electrochemical plants more feasible in comparison to thermochemical configurations.

Cost implications of alternative biomass feedstock price are also investigated while keeping the cost of electricity and carbon dioxide constant at 14 €/GJ (50 €/MWh) and 40 €/t, respectively (see Fig. 8). The LCOFs of electrochemical plants

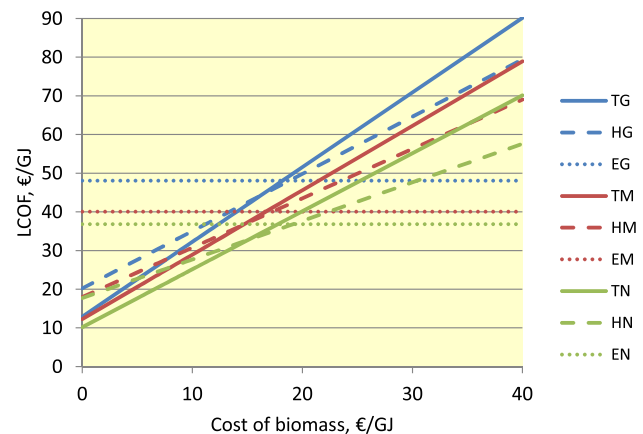


Fig. 8 – Levelised cost of fuel production (LCOF) for the examined plant configurations as a function of the biomass price.

are naturally unsensitive to changes in the cost of biomass feedstock. Hybrid and thermochemical plants are almost equally sensitive, although the slopes for thermochemical plants are steeper. When the price of biomass feedstock changes by 1 €/GJ, it causes a change in the LCOF that is, on average, 1.7 €/GJ for thermochemical and 1.3 €/GJ for hybrid plants. According to the results, purely thermochemical plants have lower production costs than corresponding hybrid plants producing the same fuel when the price of biomass stays under 14 €/GJ (50 €/MWh). For purely electrochemical configurations this threshold biomass feedstock price is about 17 €/GJ (61 €/MWh).

5.7. Preconditions for electrolytic hydrogen

The threshold electricity price that makes hybrid configurations more feasible than thermochemical plants was found to be 6 €/GJ. For purely electrochemical plants this value was found to be about 4 €/GJ. Using these electricity prices, the production cost of hydrogen is calculated to be 14 €/GJ for hybrid plants and 11 €/GJ for electrochemical plants. Therefore, when hydrogen is produced at a lower cost than these values, hybrid and electrochemical configurations become more feasible in comparison to thermochemical plants.

If the desire is to run the electrolysis only during times of excess renewable electricity, the impact of intermittent production to the levelised cost of hydrogen (LCOH) calls for additional analysis. This is carried out by calculating those electricity prices that maintain these threshold hydrogen prices (LCOH 14 and 11 €/GJ) at different annual operating hours, indicated by an on-stream factor (100% on-stream factor = 8766 h/a).

These results are illustrated in Fig. 9. It can be seen how the LCOHs gradually become more and more sensitive to the price of electricity as the on-stream factor becomes smaller. For both threshold values, there is a point on the axis where the average price of electrolyser feedstock (electricity) must go negative in order to maintain the fixed LCOHs as the annual operating hours continue to decrease: for the hybrid processes this happens at 40% (3530 h/a) and for electrochemical plants

²⁶ Average prices for chlorine and ammonia sectors taken from Refs. [74], based on data from Centre for European Policy Studies.

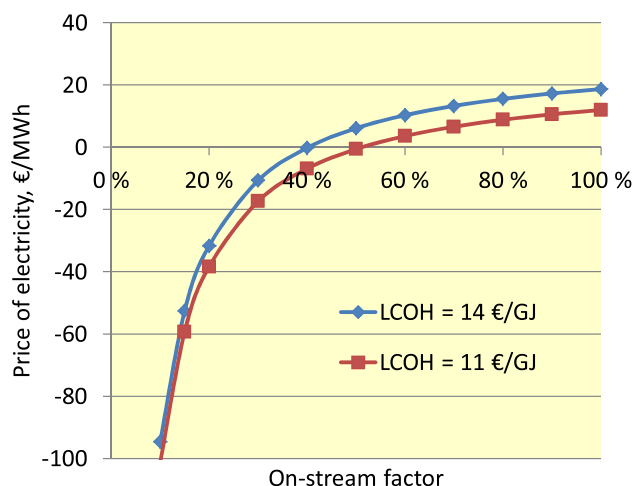


Fig. 9 – Electricity prices for an alkaline electrolysis system having a levelised cost of hydrogen (LCOH) of 14 or 11 €/GJ as a function of the on-stream factor.

at 51% (4490 h/a). For on-stream factors smaller than 20%, the LCOHs become highly sensitive to the price of electricity. For example, if the electrolyser would operate only 10% of the year (877 h) the average price of electricity would need to be –91 €/MWh and –97 €/MWh to keep LCOHs at 14 €/GJ and 11 €/GJ, respectively.

Lastly, the impact of the electrolyser's investment cost is analysed. Fig. 10 shows LCOHs as a function of the specific investment cost while keeping the price of electricity and annual operating hours constant at 14 €/GJ and 8000 h, respectively.²⁷ The results are calculated for an alkaline electrolyser having a system efficiency of 62% (LHV) and also for an 100% efficient 'ideal' electrolyser.

When the specific investment cost changes by 100 €/, it causes a change in the LCOH that is 1.1 €/GJ for an alkaline electrolyser and 0.7 €/GJ for an 'ideal' electrolyser. Somewhat surprisingly, the target hydrogen prices (14 and 11 €/GJ) can not be reached even with an 100% efficient electrolyser system, running 8000 h annually and having zero investment cost.

6. Conclusions

A detailed and transparent analysis on the performance and costs of producing synthetic fuels from biomass, carbon dioxide and electricity has been presented, based on technologies that are either commercially available or at the very least successfully demonstrated at precommercial scale. The overall economics were evaluated from the perspective of a synthetic fuel producer in terms of euros (€) per gigajoule (GJ). The costs were:

- 18 €/GJ (natural gas), 21 €/GJ (methanol) and 23 €/GJ (gasoline) for purely thermochemical plants;

²⁷ The value of byproduct oxygen is not considered as it was already included when the target LCOH values were calculated.

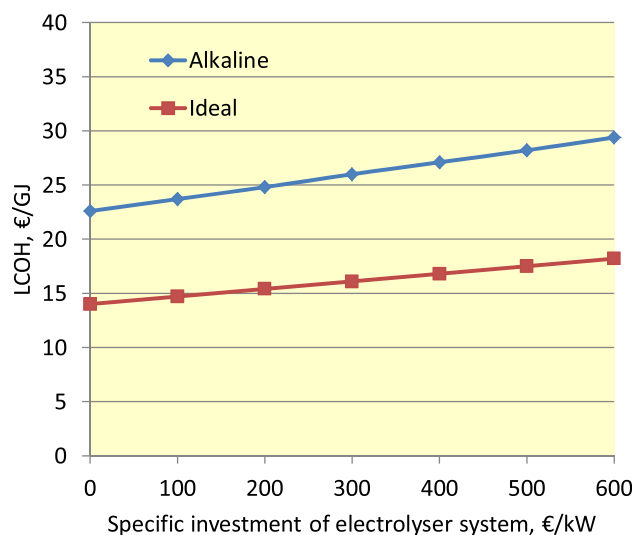


Fig. 10 – Levelised cost of hydrogen production (LCOF) for alkaline (system efficiency 62% (LHV)) and ideal (100%) electrolyser systems as a function of specific investment cost.

- 23 €/GJ (natural gas), 24 €/GJ (methanol) and 28 €/GJ (gasoline) for hybrid plants;
- 37 €/GJ (natural gas), 40 €/GJ (methanol) and 48 €/GJ (gasoline) for electrochemical plants.

For a given end-product, the lowest costs were associated with thermochemical production from forest residues, while the highest costs were associated with electrochemical production from carbon dioxide and electricity. The hybrid plants were found capable of producing fuels at a lower cost than purely electrochemical plants, but not lower than purely thermochemical plants. For all production routes, natural gas was the cheapest fuel to produce, followed by methanol and gasoline. However, out of the examined options, gasoline is the only fuel that can be readily consumed in the current transportation infrastructure without limitations, and costs related to the required modifications in vehicles and/or fuel distribution infrastructure were not included in the costs of natural gas and methanol.

It should be noted that the production costs for all the examined configurations are higher than the current price of fossil gasoline of about 15 €/GJ²⁸ and sustained subsidies either in the form of investment support, mandates or carbon price is required to make these processes economically feasible.

An analysis of the impact of feedstock prices to fuel production costs suggest that the above findings are generally robust. The following results were derived from sensitivity analysis:

- Electricity at the cost of 4 €/GJ (14 €/MWh) or lower is required to make electrochemical plants more feasible

²⁸ Based on \$100/bbl crude, \$14/bbl refining margin and 1.3 €/€/\$ exchange rate.

than corresponding thermochemical plants; for hybrid plants this threshold electricity price is 6 €/GJ (22 €/MWh).

- Zero cost CO₂ feedstock is not enough to make electrochemical plants more feasible than thermochemical or hybrid plants.
- Thermochemical plants are more feasible than corresponding hybrid plants when the cost of biomass feedstock is less than 14 €/GJ (50 €/MWh). For purely electrochemical plants this threshold biomass feedstock price is 17 €/GJ (61 €/MWh).

An analysis of the economic preconditions of using electrolytic hydrogen for synthetic biofuels production attests for the importance of long annual operating hours associated with continuous availability of low-cost renewable electricity:

- If electrochemical plants operate less than 4500 h annually, negative electricity prices are needed to make them more feasible than corresponding thermochemical plants. For hybrid plants this threshold value is 3500 h/a.
- Continuously operating electrochemical plants using an electrolyser system having zero investment cost and 100% efficiency are less feasible than thermochemical plants under the assumed performance and cost parameters.

According to these findings, power-to-fuels concepts, despite their exciting technical potential, seem to be characterised by high overall costs under practical financial assumptions. From the perspective of a synthetic fuel producer, thermochemical route from biomass was found to be substantially more feasible than electrochemical route from carbon dioxide and electricity. However, additional analysis should be carried out to explore whether feasibility of the hybrid process could be further improved by deeper integration or more sophisticated operational strategies.

Acknowledgements

The research leading to these findings was funded by VTT Technical Research Centre of Finland and Tekes the Finnish Funding Agency for Innovation.

Appendix A

In order to enable the reader to reproduce the performance and cost analysis presented in this paper, detailed process design parameters are given in Table A.11

Table A.11: Process design parameters for the examined plant configurations.

Item	Design parameters	Notes
Air separation unit	Oxygen delivered from ASU at 1.05 bar pressure. Oxygen product (mol-%): O ₂ = 99.5%, N ₂ = 0.5%, Ar = 0%. Power consumption 263 kWh/tonO ₂ .	a
Feedstock preparation and handling	Feeding screw power consumption 7 kJ/kg biomass. Lock-hopper inert gas consumption: 0.07642 Nm ³ /kg _{BIOMASS} for a double lock-hopper system that uses purge gas from LH to partly pressurise another LH. For a single lock-hopper system inert gas consumption 50% higher.	b
Atmospheric band conveyor dryer	Biomass moisture: inlet 50 wt-%, outlet 15 wt-%, hot water: T _{IN} = 90 °C, T _{OUT} = 60 °C, steam: 1 bar, 100 °C heat consumption 1300 kWh/tonH ₂ O _{EVAP.} , power consumption 32 kWh/ton _{DRYBIOMASS} .	c
Pressurised circulating fluidised-bed steam/O ₂ gasifier	Heat loss = 1% of biomass LHV. Δp = −0.2 bar. Carbon conversion: 98%. Modelled in two steps with RStoic and RGibbs using Redlich-Kwong-Soave equation of state with Boston-Mathias modification (RKS-BM). Hydrocarbon formation (kmol/kg of fuel volatiles): CH ₄ = 6.7826, C ₂ H ₄ = 0.4743, C ₂ H = 0.2265, C ₆ H ₆ = 0.2764. Tars modelled as naphthalene: C ₁₀ H ₈ = 0.0671, All fuel nitrogen converted to NH ₃ . All other components assumed to be in simultaneous phase and chemical equilibrium.	d, e
Ceramic hot-gas filter	Δp = −0.2 bar. Inlet temperature 550 °C.	e
Catalytic autothermal partial oxidation reformer	Modelled as RGibbs using Redlich-Kwong-Soave equation of state with Boston-Mathias modification (RKS-BM). Phase and chemical equilibrium conversion for C ₂ + and tar. Ammonia conversion restricted to 50%. Outlet temperature and CH ₄ conversion: 957 °C & 95% or 850 °C & 35% depending on the case investigated. Δp = −0.2 bar.	d, e
Sour shift	T _{OUT} = 404 °C, Steam/CO = 1.8 mol/mol, Δp = −0.2 bar. Modelled as REquil using Redlich-Kwong-Soave equation of state with Boston-Mathias modification (RKS-BM). Equilibrium reactions: CO + H ₂ O = CO ₂ + H ₂ , T _{APPR.} = 10 K. COS + H ₂ O = CO ₂ + H ₂ S, T _{APPR.} = 0 K. HCN + H ₂ O = CO + NH ₃ , T _{APPR.} = 10 K.	f, e
Scrubber	Scrubbing liquid: water. T _{INLET} 200 °C. Two-step cooling: T ¹ _{OUT} = 60 °C, T ² _{OUT} = 30 °C. Complete ammonia removal. Modelled as Flash using Soave-Redlich-Kwong (SRK) equation of state model.	e

– (continued)		
Item	Design parameters	Notes
Rectisol acid gas removal	100% H ₂ S capture, for CO ₂ capture level see case designs. Utilities: Electricity (other than for refrigeration) = 1900 kJ/kmol (CO ₂ +H ₂ S); Refrigeration 3× duty needed to cause –12 K temperature change in the syngas; 5 bar steam = 6.97 kg/kmol (H ₂ S + CO ₂).	g
Low-pressure methanol	T _{REACTION} = 260 °C, P _{MAKE-UP} = 80 bar, Δp = –5 bar, Boiling-water reactor modelled with REquil using Soave-Redlich-Kwong equation of state (SRK). Equilibrium reactions: CO + 2H ₂ = CH ₄ O, T _{APPR.} = 10 K; CO ₂ + 3H ₂ = CH ₄ O + H ₂ O, T _{APPR.} = 10 K.	e
High temperature methanation	Six adiabatic fixed-bed reactors connected in series and equipped with intercoolers. Pressure at system inlet = 15 bar, pressure at system outlet 11 bar. T _{INPUT} to reactors 300 °C. T _{OUTPUT} from the first reactor restricted to 700 °C with steam dilution. Gas dried before feeding to last reactor. Syngas conversion to methane ≥99.5%. Equilibrium reactions: CO + 3H ₂ = CH ₄ + H ₂ O, T _{APPR.} = 20 K; CO ₂ + 4H ₂ = CH ₄ + 2H ₂ O, T _{APPR.} = 20 K. Reactors modelled as REquils using Soave-Redlich-Kwong (SRK) equation of state model.	e
Methanol-to-Gasoline	DME reactor: T _{IN} = 297 °C, T _{OUT} = 407 °C, P _{IN} = 23 bar, Δp = –1 bar, Boiling-water reactor modelled with REquil using Soave-Redlich-Kwong equation of state (SRK). Equilibrium reaction: 2CH ₄ O = C ₂ H ₆ O + H ₂ O, T _{APPR.} = 30 K. Gasoline reactor: T _{REACTOR} = 400 °C, P _{IN} = 22 bar, Δp = –1 bar, Modelled as REquil using Soave-Redlich-Kwong equation of state (SRK). Relative mass yields from 1 ton of raw product in the refining area are 880 kg of gasoline blendstock, 100 kg of LPG and 20 kg of purge gas.	h
Alkaline electrolysis	H ₂ and O ₂ purity 100%. Both delivered at atmospheric pressure and 25 °C, Electrolyser system efficiency = 62% (LHV).	e, i
Auxiliary boiler	Modelled as RStoic, Δp = –0.1 bar, Lambda = 1.20, Air preheat to 250 °C with fluegas	e
Heat exchangers	Δp/p = 2%; ΔT _{MIN} = 15 °C (gas-liq), 30 °C (gas-gas). Heat loss = 1% of heat transferred.	g
Heat recovery & Steam system	Flue gas T _{OUT} = 150 °C, feed water pressure 110 bar, steam drum blowdowns: 2% of inlet flow, Deaerator T _{OUT} = 120 °C.	e
Steam turbine	Inlet steam parameters: 93.5 bar, 500 °C; Extraction steam parameters: HP = 25 bar, 330 °C; LP = 5 bar, 179 °C; η _{ISENTROPIC} = 0.78, η _{GENERATOR} = 0.97, η _{MECHANICAL} = 0.98.	c,e,j
Compressors	Stage pressure ratio <2, η _{POLYTROPIC} = 0.85, η _{DRIVER} = 0.92, η _{MECHANICAL} = 0.98.	k
Multistage compressors (>4.5 kg/s)	Stage pressure ratio <2, η _{POLYTROPIC} = 0.87, η _{DRIVER} = 0.92, η _{MECHANICAL} = 0.98, T _{INTERCOOLER} = 35 °C, Δp/p _{INTERCOOLER} = 1%.	l
Multistage compressors (<4.5 kg/s) ⁶	Stage pressure ratio <2, η _{POLYTROPIC} = 0.85, η _{DRIVER} = 0.90, η _{MECHANICAL} = 0.98, T _{INTERCOOLER} = 35 °C, Δp/p _{INTERCOOLER} = 1%	l
Pumps	H _{HYDRAULIC} = 0.75, η _{DRIVER} = 0.90.	k

a – Taken from Smith et al. [80].
b – Taken from Swanson et al. [81]. The original value in the reference was given for bagasse (160 kg/m³), which is here fitted for forest residues (293 kg/m³) assuming that LH is filled with feedstock up to 90%.
c – Based on personal communication with Andras Horvath, Carbona-Andritz, May 15th 2012.
d – Modelling principles taken from Refs. [82] and [35].
e – Operating parameters chosen by author.
f – Outlet temperature and steam/CO ratio based on personal communication with Wolfgang Kaltner, Süd-Chemie AG, July 9th, 2012.
g – Taken from Liu et al. [73].
h – Taken from Larson et al. [49]. For MTG reactor yield structure, see section 3.4.
i – System efficiency calculated based on information taken from Ivy [67].
j – Based on personal communication with Reijo Kallio, ÅF-Consult, October 2012.
k – Taken from Chiesa et al. [83].
l – Taken from Glassman [84].

REFERENCES

- [1] IPCC. Summary for policymakers. In: Pachauri R, Reisinger A, editors. Climate change 2007: synthesis report. Geneva: Switzerland; 2007.
- [2] CO₂ emissions from fuel combustion – highlights. 9 rue de la Federation 75739 Paris Cedex 15, France: International Energy Agency; 2013.
- [3] Breeze P. Coping with carbon: a near-term strategy to limit carbon dioxide emissions from power stations, Philosophical Transactions of the Royal Society A: mathematical. Phys Eng Sci 2008;366(1882):3891–900. <http://dx.doi.org/10.1098/rsta.2008.0113>.
- [4] Zeman FS, Keith DW. Carbon neutral hydrocarbons, Philosophical Transactions of the Royal Society A: mathematical. Phys Eng Sci 2008;366(1882):3901–18.
- [5] Kreutz T, Larson E, Williams R, Liu G. Fischer–Tropsch fuels from coal and biomass. Pittsburgh, PA, USA. In: Proceedings

- of the 25th Annual International Pittsburgh Coal Conference; 2008.
- [6] G. Liu, E. D. Larson, R. H. Williams, T. G. Kreutz, X. Guo. Making Fischer-Tropsch fuels and electricity from coal and biomass: performance and cost analysis, *Energy Fuels*;25(1). doi:10.1021/ef101184e.
 - [7] Van Bibber L, Shuster E, Haslbeck J, Rutkowski M, Olsen S, Kramer S. Baseline technical and economic assessment of a commercial scale Fischer–Tropsch liquids facility. Tech Rep DOE/NETL-2007/1260. National Energy Technology Laboratory; 2007.
 - [8] Hamelinck C, Faaij A, den Uil H, Boerrigter H. Production of FT transportation fuels from biomass; technical options, process analysis and optimisation, and development potential. *Energy* 2004;29(11):1743–71.
 - [9] McKeough P, Kurkela E. Process evaluations and design studies in the UCG project 2004–2007. VTT Research Notes 2434. Technical Research Centre of Finland, VTT; 2008.
 - [10] Landälv I. Demonstration plants for advanced biofuels production based on thermochemical pathways – status and reflections. In: Proceedings of the 21st European Biomass Conference and Exhibition, Copenhagen, Denmark; 2013.
 - [11] Audi unveils e-gas project, Green car congress. May 13, 2011. bit.ly/1qHXLke.
 - [12] Rath L. Cost and performance baseline for fossil energy plants volume 2: coal to synthetic natural gas and ammonia. Final Report, July 2011 DOE/NETL-2010/1402. U.S. National Energy Technology Laboratory; 2011.
 - [13] Appl M. In: More Alexander, editor. Ammonia, methanol, hydrogen, carbon monoxide – Modern production technologies, Nitrogen. 31 Mount Pleasant, London, WC1X 0AD England: CRU Publishing Ltd; 1997, ISBN 1 873387 26 1.
 - [14] Dybkjaer I, Hansen JB. Large-scale production of alternative synthetic fuels from natural gas. In: de Pontes M, Espinoza R, Nicolaidis C, Scholtz J, Scurrell M, editors. Natural gas conversion IV. Studies in surface science and catalysis, 107. Elsevier; 1997. p. 99–116. [http://dx.doi.org/10.1016/S0167-2991\(97\)80322-9](http://dx.doi.org/10.1016/S0167-2991(97)80322-9).
 - [15] Penick J, Lee W, Maziuk J. Development of the methanol-to-gasoline process. Chemical reaction engineering – plenary lectures, ACS Symposium Series, 226; 1983. p. 19–48. Ch. 3.
 - [16] German biofuel firm Choren declares insolvency. Reuters; July 8, 2011. bit.ly/1nZYaus.
 - [17] Range fuels cellulosic ethanol plant fails, U.S. pulls plug. Bloomberg; December 3, 2011. bloom.bg/1lp8eJB.
 - [18] Oil Neste. Stora Enso shelve plans for renewable diesel plant. Biomass Magazine; August 22, 2012. bit.ly/1mUBfPz.
 - [19] Report from the bio-DME plant in Piteå. ChemrecNews; May, 2012. bit.ly/1u9ailN.
 - [20] Haldor Topsøe and partners produce biobased gasoline. Biomass Magazine; June 26, 2013. bit.ly/1mUBvOU.
 - [21] M. Steinberg, Electrolytic synthesis of methanol from CO₂, U.S. patent 3959094, United States Energy Research and Development Administration (1976).
 - [22] J. Lewis, A. Martin, Method for obtaining carbon dioxide from the atmosphere and for production of fuels, U.S. patent 4140602, Texas Gas Transmission Corporation (1979).
 - [23] M. Steinberg, Synthetic carbonaceous fuels and feedstocks, U.S. patent 4197421, United States Energy Research and Development Administration (1980).
 - [24] M. Corbett, S. Salina, Production of synthetic hydrocarbons from air, water and low cost electrical power, U.S. patent 4282187, Grumman Aerospace Corporation (1981).
 - [25] Graves C, Ebbesen SD, Mogensen M, Lackner KS. Sustainable hydrocarbon fuels by recycling CO₂ and H₂O with renewable or nuclear energy. *Renew Sustain Energy Rev* 2011;15(1):1–23. <http://dx.doi.org/10.1016/j.rser.2010.07.014>.
 - [26] Inui T. Highly effective conversion of carbon dioxide to valuable compounds on composite catalysts. *Catal Today* 1996;29(14):329–37. [http://dx.doi.org/10.1016/0920-5861\(95\)00300-2](http://dx.doi.org/10.1016/0920-5861(95)00300-2). Second Japan–EC Joint Workshop on the Frontiers of Catalytic Science and Technology for Energy, Environment and Risk Prevention.
 - [27] Kieffer R, Fujiwara M, Udron L, Souma Y. Hydrogenation of CO and CO₂ toward methanol, alcohols and hydrocarbons on promoted copper-rare earth oxides catalysts. *Catal Today* 1997;36(1):15–24. [http://dx.doi.org/10.1016/S0920-5861\(96\)00191-5](http://dx.doi.org/10.1016/S0920-5861(96)00191-5). Copper, Silver and Gold in Catalysis.
 - [28] Saeidi S, Amin NAS, Rahimpour MR. Hydrogenation of CO₂ to value-added products – a review and potential future developments. *J CO₂ Util* 2014;5(0):66–81. <http://dx.doi.org/10.1016/j.jcou.2013.12.005>.
 - [29] Ouellette N, Rogner H-H, Scott D. Hydrogen from remote excess hydroelectricity. part ii: hydrogen peroxide or biomethanol. *Int J Hydrogen Energy* 1995;20(11):873–80. [http://dx.doi.org/10.1016/0360-3199\(95\)00018-9](http://dx.doi.org/10.1016/0360-3199(95)00018-9).
 - [30] Cifre PG, Badr O. Renewable hydrogen utilisation for the production of methanol. *Energy Convers Manag* 2007;48(2):519–27. <http://dx.doi.org/10.1016/j.enconman.2006.06.011>.
 - [31] Mignard D, Pritchard C. On the use of electrolytic hydrogen from variable renewable energies for the enhanced conversion of biomass to fuels. *Chem Eng Res Des* 2008;86(5):473–87. <http://dx.doi.org/10.1016/j.cherd.2007.12.008>.
 - [32] Kurkela E, Kurkela M, Hiltunen I. The effects of wood particle size and different process variables on the performance of steam-oxygen blown circulating fluidized-bed gasifier. *Environ Prog Sustain Energy* 2014;33(3):681–7. <http://dx.doi.org/10.1002/ep.12003>.
 - [33] Simell P, Hannula I, Tuomi S, Nieminen M, Kurkela E, Hiltunen I, et al. Clean syngas from biomass – process development and concept assessment. Biomass Convers Biorefinery 2014;1–14. <http://dx.doi.org/10.1007/s13399-014-0121-y>.
 - [34] Kurkela E, Simell P, McKeough P, Kurkela M. Production of synthesis gas and clean fuel gas [Synthesikaasun ja puhtaan polttoakaasun valmistus]. VTT Publications 682, Technical Research Centre of Finland, VTT; 2008.
 - [35] Hannula I, Kurkela E. A parametric modelling study for pressurised steam/O₂-blown fluidised-bed gasification of wood with catalytic reforming. *Biomass Bioenergy* 2012;38(0):58–67. <http://dx.doi.org/10.1016/j.biombioe.2011.02.045>. Overcoming Barriers to Bioenergy: Outcomes of the Bioenergy Network of Excellence 2003–2009.
 - [36] Hannula I, Kurkela E. Liquid transportation fuels via large-scale fluidised-bed gasification of lignocellulosic biomass. VTT Technology 91, Technical Research Centre of Finland; 2013.
 - [37] Sabatier P, Senderens J. Nouvelles synthèses du methane. *Acad Sci* 1902;314:514–6.
 - [38] Kopyscinski J, Schildhauer TJ, Biollaz SM. Production of synthetic natural gas (SNG) from coal and dry biomass – a technology review from 1950 to 2009. *Fuel* 2010;89(8):1763–83.
 - [39] From solid fuels to substitute natural gas (SNG) using TREMP, Tech. rep. Haldor Topsøe; 2009.
 - [40] Hansen JB, Hojlund Nielsen PE. Methanol synthesis. In: Handbook of heterogeneous catalysis. Wiley-VCH Verlag GmbH & Co. KGaA; 2008. <http://dx.doi.org/10.1002/9783527610044.hetc0148>.
 - [41] Tijm P, Waller F, Brown D. Methanol technology developments for the new millennium. *Appl Catal A Gen* 2001;221(1–2):275–82. [http://dx.doi.org/10.1016/S0926-860X\(01\)00805-5](http://dx.doi.org/10.1016/S0926-860X(01)00805-5). Hoelderich Special Issue.

- [42] Yurchak S. Development of Mobil's fixed-bed methanol-to-gasoline (MTG) process. *Stud Surf Sci Catal* 1988;36:251–72.
- [43] P. Nielsen, F. Joensen, J. Hansen, E. Sørensen, J. Madsen, R. Mabrouk, Process for the preparation of hydrocarbons from oxygenates, Patent US 8067474 B2, Haldor Topsøe A/S (2011).
- [44] E. Jörn, J. Røstrup-Nielsen, Process for the preparation of hydrocarbons, Patent US 4481305, Haldor Topsøe A/S (1984).
- [45] Chang CD, Silvestri AJ. The conversion of methanol and other O-compounds to hydrocarbons over zeolite catalysts. *J Catal* 1977;47(2):249–59.
- [46] Chang C. Hydrocarbons from methanol. *Catal Rev* 1983;25(1):1–118.
- [47] Wiley critical content – petroleum technology, vol. 1–2, John Wiley & Sons, 2007.
- [48] Kam A, Schreiner M, Yurchak S. Mobil methanol-to-gasoline process. In: Meyers RA, editor. *Handbook of synfuels technology*. McGraw-Hill.; 1984.
- [49] Larson E, Williams R, Kreutz T, Hannula I, Lanzini A, Liu G. Energy, environmental, and economic analyses of design concepts for the co-production of fuels and chemicals with electricity via co-gasification of coal and biomass. Final report under contract DEFE0005373 to the National Energy Technology Laboratory, US Department of Energy, Princeton University. 2012.
- [50] Barker E, Begovich J, Clinton J, Johnson P. Aspen modeling of the tri-state indirect liquefaction process. contract no. W-7405-eng-6. Oak Ridge National Laboratory; 1983.
- [51] Schreiner M. Research guidance studies to assess gasoline from coal by methanol-to-gasoline and sasol-type Fischer–Tropsch technologies. Final report to doe under contract no. ef-77-c-01-2447. 1978.
- [52] Keith D, Ha-Duong M, Stolaroff J. Climate strategy with CO₂ capture from the air. *Clim Change* 2006;74(1–3):17–45. <http://dx.doi.org/10.1007/s10584-005-9026-x>.
- [53] Martin F, Kubic W. Green freedom – a concept for producing carbon-neutral synthetic fuels and chemicals, Overview. Los Alamos National Laboratory; 2007.
- [54] Hamilton MR, Herzog HJ, Parsons JE. Cost and U.S. public policy for new coal power plants with carbon capture and sequestration. *Energy Procedia* 2009;1(1):4487–94. <http://dx.doi.org/10.1016/j.egypro.2009.02.266>. Proceedings of the 9th International Conference on Greenhouse Gas Control Technologies (GHGT-9), 1620 November 2008, Washington DC, USA.
- [55] Rubin ES, Chen C, Rao AB. Cost and performance of fossil fuel power plants with CO₂ capture and storage. *Energy Policy* 2007;35(9):4444–54. <http://dx.doi.org/10.1016/j.enpol.2007.03.009>.
- [56] Metz B, Davidson O, de Coninck H, Loos M, Meyer L, editors. IPCC special report on carbon dioxide capture and storage. Prepared by working Group III of the Intergovernmental Panel on Climate change. Cambridge, United Kingdom and New York, NY, USA: Cambridge University Press; 2005.
- [57] Song C. Global challenges and strategies for control, conversion and utilization of CO₂ for sustainable development involving energy, catalysis, adsorption and chemical processing. *Catal Today* 2006;115(1–4):2–32.
- [58] Inui T, Takeguchi T. Effective conversion of carbon dioxide and hydrogen to hydrocarbons. *Catal Today* 1991;10(1):95–106. [http://dx.doi.org/10.1016/0920-5861\(91\)80077-M](http://dx.doi.org/10.1016/0920-5861(91)80077-M).
- [59] Inui T, Kitagawa K, Takeguchi T, Hagiwara T, Makino Y. Hydrogenation of carbon dioxide to C₁–C₇ hydrocarbons via methanol on composite catalysts. *Appl Catal A Gen* 1993;94(1):31–44. [http://dx.doi.org/10.1016/0926-860X\(93\)80043-P](http://dx.doi.org/10.1016/0926-860X(93)80043-P).
- [60] Sahibzada M, Chadwick D, Metcalfe I. Hydrogenation of carbon dioxide to methanol over palladium-promoted Cu/ZnO/Al₂O₃ catalysts. *Catal Today* 1996;29(14):367–72. [http://dx.doi.org/10.1016/0920-5861\(95\)00306-1](http://dx.doi.org/10.1016/0920-5861(95)00306-1). second Japan-EC Joint Workshop on the Frontiers of Catalytic Science and Technology for Energy, Environment and Risk Prevention.
- [61] Ushikoshi K, Moria K, Watanabe T, Takeuchi M, Saito M. A 50 kg/day class test plant for methanol synthesis from CO₂ and H₂. In: Inui T, Anpo M, Izui K, Yanagida S, Yamaguchi T, editors. *Advances in Chemical Conversions for Mitigating Carbon Dioxide* Proceedings of the Fourth International Conference on Carbon Dioxide Utilization. Studies in Surface Science and Catalysis, vol. 114. Elsevier; 1998. p. 357–62. [http://dx.doi.org/10.1016/S0167-2991\(98\)80770-2](http://dx.doi.org/10.1016/S0167-2991(98)80770-2).
- [62] Saito M. R&D activities in Japan on methanol synthesis from CO₂ and H₂. *Catal Surv Jpn* 1998;2(2):175–84.
- [63] Amenomiya Y. Methanol synthesis from CO₂ + H₂ II. Copper-based binary and ternary catalysts. *Appl Catal* 1987;30(1):57–68. [http://dx.doi.org/10.1016/S0166-9834\(00\)81011-8](http://dx.doi.org/10.1016/S0166-9834(00)81011-8).
- [64] Bertuccioli L, Chan A, Hart D, Lehner F, Madden B, Standen E. Study on development of water electrolysis in the EU, fuel cells and hydrogen joint undertaking. 2014.
- [65] Holladay J, Hu J, King D, Wang Y. An overview of hydrogen production technologies. *Catal Today* 2009;139(4):244–60. <http://dx.doi.org/10.1016/j.cattod.2008.08.039>. Hydrogen Production – Selected papers from the Hydrogen Production Symposium at the American Chemical Society 234th National Meeting & Exposition, August 19–23, 2007, Boston, MA, USA.
- [66] Turner J, Sverdrup G, Mann MK, Maness P-C, Kroposki B, Ghirardi M, et al. Renewable hydrogen production. *Int J Energy Res* 2008;32(5):379–407. <http://dx.doi.org/10.1002/er.1372>.
- [67] Ivy J. Summary of electrolytic hydrogen production – milestone completion report. Tech. Rep. NREL/MP-560–36734. National Renewable Energy Laboratory; 2004.
- [68] Merrow E, Phillips K, Myers C. Understanding cost growth and performance shortfalls in pioneer process plants. Tech. Rep. RAND/R-2569-DOE. Santa Monica, USA: RAND Corporation; 1981.
- [69] Kaukolämpötilasto, statistics in Finnish. Energiatietokes; 2012. bit.ly/1qjFidI.
- [70] Black J. Cost and performance baseline for fossil energy plants volume 1: bituminous coal and natural gas to electricity. Revision 2a, September 2013 DOE/NETL-2010/1397. U.S. National Energy Technology Laboratory; 2007.
- [71] Wan V. Methanol to olefins. Report no. 261. SRI Consulting; 2007.
- [72] Kreutz T, Williams R, Consonni S, Chiesa P. Co-production of hydrogen, electricity and CO₂ from coal with commercially ready technology. Part B: economic analysis. *Int J Hydrogen Energy* 2005;30(7):769–84. <http://dx.doi.org/10.1016/j.ijhydene.2004.08.001>.
- [73] G. Liu, E. D. Larson, R. H. Williams, T. G. Kreutz, X. Guo, Supporting information for making Fischer–Tropsch fuels and electricity from coal and biomass: performance and cost analysis, *Energy Fuels* 25(1). doi:10.1021/ef101184e.
- [74] Energy prices and costs report, European Commission staff working document. 2014. bit.ly/1tANeta.
- [75] Wilen C, Moilanen A, Kurkela E. Biomass feedstock analyses. VTT Publications 282, Technical Research Centre of Finland, VTT; 1996.
- [76] Larson ED, Jin H, Celik FE. Large-scale gasification-based coproduction of fuels and electricity from switchgrass. *Biofuels Bioprod Biorefining* 2009;3(2):174–94. <http://dx.doi.org/10.1002/bbb.137>.

-
- [77] Project co-funded by the European Commission within the Sixth Framework Programme project no. 019761 Bigpower deliverable 71: Finnish case study report. Carbona, Inc.; 2009.
- [78] Floch P, Gabriel S, Mansilla C, Werkoff F. On the production of hydrogen via alkaline electrolysis during off-peak periods. *Int J Hydrogen Energy* 2007;32(18):4641–7. <http://dx.doi.org/10.1016/j.ijhydene.2007.07.033>.
- [79] Ekbom T, Lindblom M, Berglin N, Ahlvik P. Technical and commercial feasibility study of black liquor gasification with methanol/DME production as motor fuels for automotive uses – BLGMF. Tech. Rep. Contract No. 4.1030/Z/01–087/2001. Nykomb Synergetics; 2003.
- [80] Smith A, Klosek J. A review of air separation technologies and their integration with energy conversion processes. *Fuel Process Technol* 2001;70(2):115–34. [http://dx.doi.org/10.1016/S0378-3820\(01\)00131-X](http://dx.doi.org/10.1016/S0378-3820(01)00131-X).
- [81] Swanson M, Musich M, Schmidt D, Schultz J. Feed system innovation for gasification of locally economical alternative fuels (FIGLEAF) final report, Tech. rep.. U.S. Department of Energy; 2002.
- [82] Hannula I, Kurkela E. A semi-empirical model for pressurised air-blown fluidised-bed gasification of biomass. *Bioresour Technol* 2010;101(12).
- [83] Chiesa P, Consonni S, Kreutz T, Williams R. Co-production of hydrogen, electricity and CO₂ from coal with commercially ready technology. Part A: performance and emissions. *Int J Hydrogen Energy* 2005;30(7):747–67. <http://dx.doi.org/10.1016/j.ijhydene.2004.08.002>.
- [84] Glassman A. Users manual for updated computer code for axial-flow compressor conceptual design, Tech. rep.. Toledo, Ohio: University of Toledo; 1992.