

# Supplementary material

## Electrification of biorefinery concepts for improved productivity – yield, economic and GHG performances

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## 1 PROCESS MODELLING

A brief description of the process modelling including main assumptions, modelling approach and simplified process block diagrams are presented here.

### 1.1 ELECTROLYSIS

Electrolysis is not a separate production track, but it is described here as part of supplementary material since the technology is used as an electrification option in many of the production tracks. Two different electrolysis technology, proton exchange membrane (PEMEC) and high-temperature solid oxide (SOEC) are considered for integration to the production process.

Figure S 1 illustrates the simplified system schematic of both electrolyzer systems. All systems presented in this report use the same system configuration, while the size of the electrolysis is set to match the given hydrogen requirement. The required heat for increasing the water feed temperature to operating temperature is provided firstly from internal heat integration in the electrolysis system, or with the biofuel production plant where possible. Electricity is used for the final high-temperature heating for SOEC. Internal heat integration is indeed more vital for SOEC as this system operates at higher temperatures.

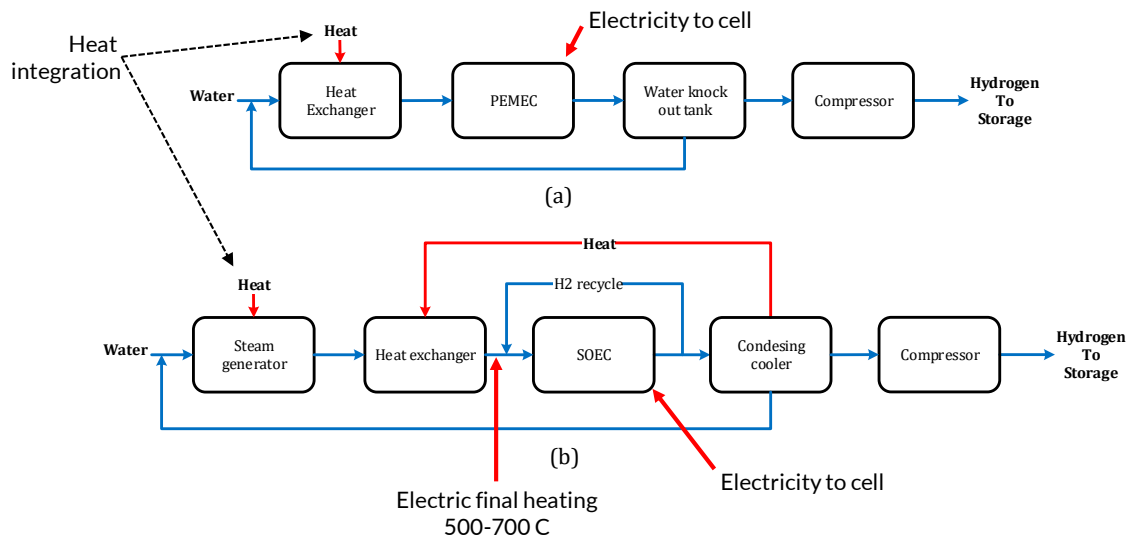


Figure S 1. Simplified system schematic of a) PEMEC and b) SOEC

Electrolyzer systems are modeled using Aspen Plus, and values of key operating parameters are summarized in Table S 1. Built in blocks from the Aspen Plus library was used to model heat exchangers as well as compressors. However, there is no readily available block to represent electrolyzer unit itself. In this case, a stoichiometric reactor linked to user defined calculator blocks was used to introduce the electrochemical reactions as well as electrical power requirement and other operational parameters such as voltage. The table below shows the input parameters used in the model development and few of key results based on kg H<sub>2</sub> that is produced by each system.

Table S 1. Electrolysis system assumptions and values for key parameters

Parameter	Unit	Value	
		SOEC [1]	PEMEC [2]
Temperature	C	700	70
Pressure	bar	1	1
Voltage	V	1.6	2
Utilization Factor	---	0.7	0.7
Sweep Air*	---	None	None
Hydrogen Recirculation	%	14	None
Water Consumption	kg/kg <sub>H2</sub>	8.97	8.97
Oxygen Production	kg/kg <sub>H2</sub>	7.94	7.94
Electrolysis Electrical Demand	GWh/kg <sub>H2</sub>	43	53
Hydrogen Compression 40 bar	GWh/kg <sub>H2</sub>	2.9	2.4
Heat Input	GWh/kg <sub>H2</sub>	9	0.5
Electrical Efficiency	%	77.4	62.7
System Efficiency	%	60.4	59.3

\* Anode side in case of no immediate use of produced oxygen

## 1.2 MECHANICAL VAPOR RECOMPRESSION (MVR)

Mechanical Vapor Recompression (MVR) heat pump technology is considered for steam generation in conversion pathways with low- to medium-pressure steam deficit. MVR heat pump makes use of the latent heat in low temperature vapor streams in a process by elevating the temperature/pressure of this steam to cover demands at higher temperatures. This will drastically change the amount of electricity needed to produce steam compared to electric steam generation that does not take advantage of the waste heat available in the system since no energy needs to be supplied to change the phase of the stream. When water vapor is compressed to a higher pressure the output steam will be superheated, mainly due to the isentropic efficiency. This in practice implies that the MVR will produce additional steam since injection of water in the superheated steam is done to decrease the temperature to near-saturation.

Typical use of MVR heat pumps in industry is in thermal separation processes, e.g. distillation, evaporation and drying [3]. A block flow diagram of a simplified MVR heat pump is shown in Figure S 2. Table S 2 summarizes processes that involve MVR under electrified options including the COP and % steam avoided due to heat pump.

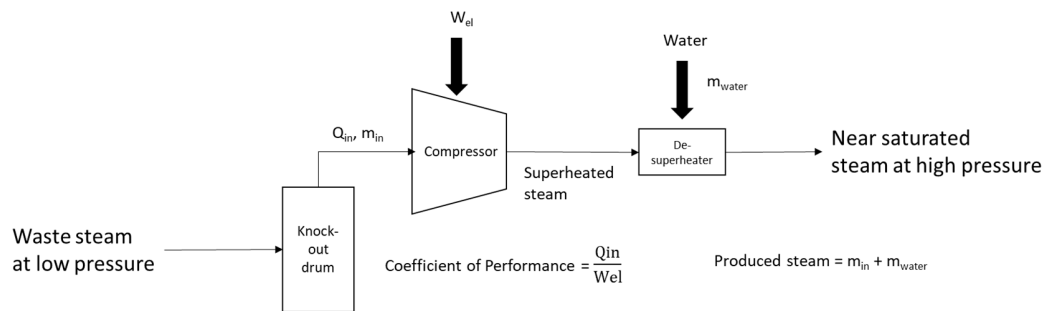


Figure S 2. Block flow diagram of a simplified MVR heat pump

Table S 2. Coefficient of performance (COP) assumptions for MVR heat pumps

Process	% steam reduction with MVR heat pump	COP
EtOH track		
Pretreatment	100	3
Distillation	100	3.5
Wood-pellet drying	50	3
E-fuels		
Carbon capture (MEA)	100	3

### 1.3 LIQUEFACTION TRACKS

#### 1.3.1 LIGNOCELLULOSIC ETHANOL (ETOH)

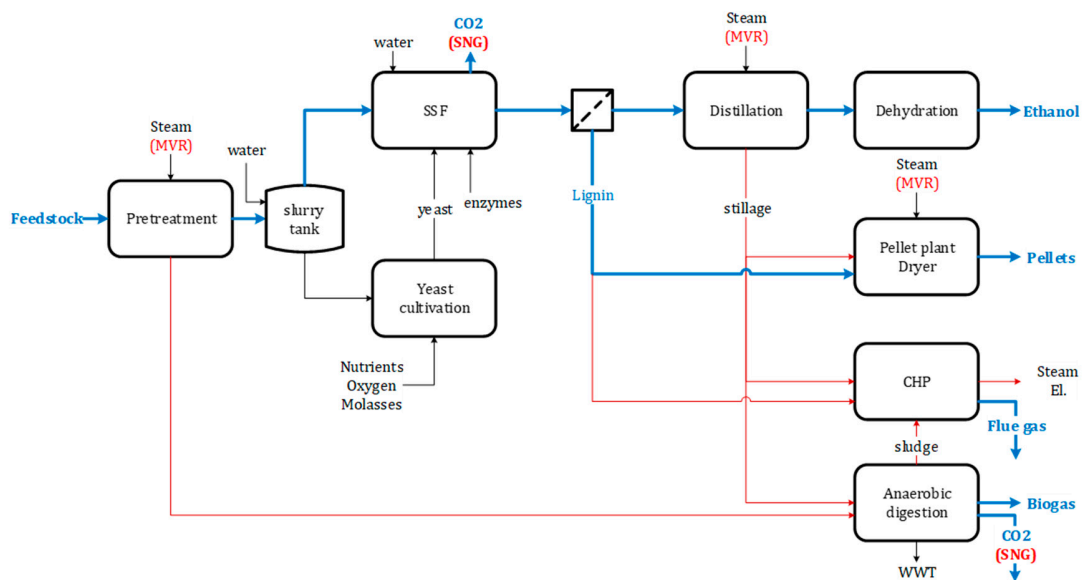


Figure S 3. Schematics of lignocellulosic ethanol (electrification options indicated in brackets)

Table S 3. Lignocellulosic ethanol modelling methodology including electrified options

Pretreatment	Pressurized mixing to 4 bar, 95°C and to 20 bar, 205°C successively. Yield reactor to decompose feedstock to fermentable sugars and water insoluble solids (WIS)
Depressurizing pretreated feedstock	Flash tanks at 4 bar and 1 bar
Fermentation	Conversion reactor at 35°C
Ethanol upgrading	Beer column at 3 bar, 15 stages, 93% vol. ethanol, reboiler run on external heat. Rectifier column – shortcut column, reboiler run on energy recovered from beer column condenser. Molecular sieve – component splitter
Anaerobic digestion (AD)	Conversion reactor
Biogas upgrading	Amine wash – component splitter
CO <sub>2</sub> compressor (electrified option)	Two stage compressors with intercooler to 40°C, polytropic efficiency 79%
Reverse water gas shift (electrified option)	Equilibrium reactor, 750°C, 30 bar
SNG synthesis (electrified option)	Three-stage equilibrium reactors in series with interstage cooling that produce HP and LP steam and a final condensing cooler to 40°C

### 1.3.2 HYDROTHERMAL LIQUEFACTION (HTL)

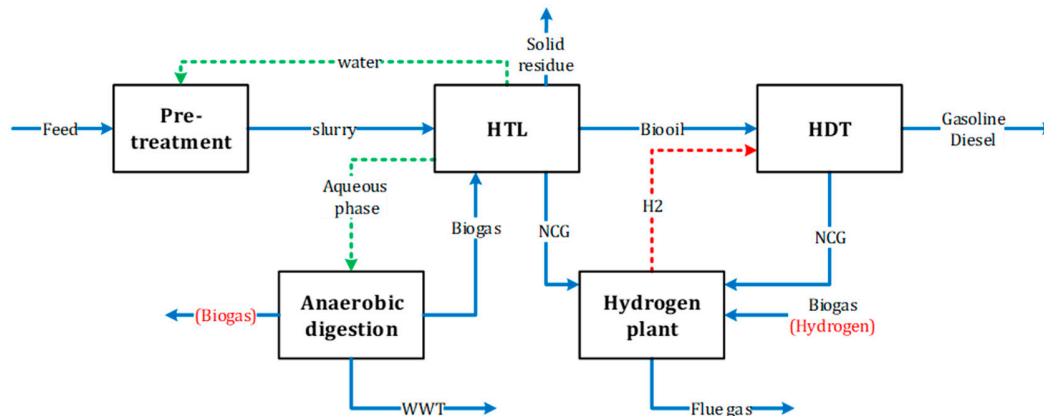


Figure S 4. Schematics of HTL of forest residue to biofuels (electrification options indicated in brackets)

Table S 4. HTL of forest residue modelling methodology

Slurry prep	Mixing tank, atmospheric
Slurry pump	Rotary pump exit pressure 208 bar, 300°C
HTL	Yield reactor at 207 bar, 300°C, fed preheated slurry at 300°C using heat recovered from HTL product effluents
H <sub>2</sub> plant	Steam reformer, exit temp. 950°C, 30 bar, modelled as eq. reactor Additional steam injection at 400°C, to favor equilibrium towards H <sub>2</sub> PSA – modelled as component splitter
HDT	Yield reactor, H <sub>2</sub> demand 0.05kg/kg biocrude
Upgrading	Distillation based on boiling point, gasoline <155°C, diesel <365°C and heavies >365°C
Anaerobic digestion (AD)	Conversion reactor
Biogas upgrading	Amine wash – component splitter
Biogas combustor	Gibb's reactor, with spec. control on flue gas O <sub>2</sub> concentration

### 1.3.3 FAST PYROLYSIS (FP)

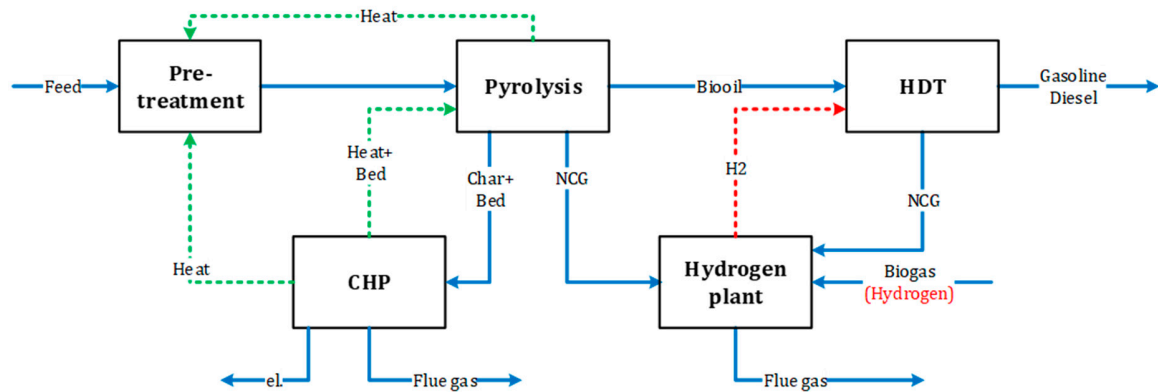


Figure S 5. Schematics of fast pyrolysis of forest residue to biofuels (electrification option indicated in bracket)

Table S 5. Fast pyrolysis and upgrading modelling approach

Pretreatment	Conveyor dryer, to 8% moisture content
Fast pyrolysis	Yield reactor, 520°C, 1.013 barg
Separation	Cyclone, separate char & bed material
Quench tower	Flash tank at 75°C cooled with recycled pyrolysis oil
H <sub>2</sub> plant	Steam reformer, exit temp. 950°C, 30 bar, modelled as eq. reactor Additional steam injection at 400°C to favor equilibrium towards H <sub>2</sub> PSA – component splitter
HDT	Yield reactor, H <sub>2</sub> demand 0.05kg/kg biocrude
Upgrading	Distillation based on boiling point, gasoline <155°C, diesel <365°C and heavies >365°C
CHP	Char combustor modelled as conversion reactor with spec. control on flue gas O <sub>2</sub> concentration. Preheated sand to pyrolysis reactor. Steam Rankine cycle, HPS at 540°C and 120 bar

#### 1.4 BLACK LIQUOR GASIFICATION (BLG)

The mass and energy balances of BLG tracks are derived based on processes modelled in UniSim Design®.

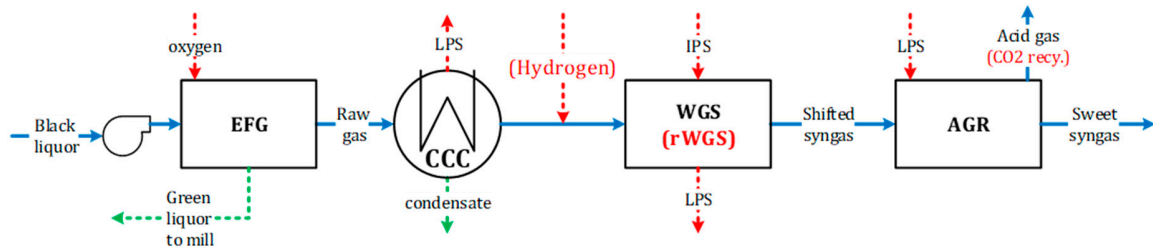


Figure S 6. Schematics of BLG up to sweet syngas (electrification options indicated in brackets)

Table S 6. BLG modeling methodology up to sweet syngas

BL pump	Centrifugal pump with outlet pressure 31 barg. Adiabatic efficiency: 75%
BLG (EFG)	Equilibrium reactor, 1050°C, 31 barg. Empirical correlation for components with poor equilibrium prediction, e.g., H <sub>2</sub> S, CH <sub>4</sub>
Counter current condenser (CCC)	Multiple heat exchangers. Flash at 80°C and 40°C
WGS reactor	Saturator 190-200°C. Equilibrium reactor ~400°C
AGR (amine wash)	Component splitter. Steam demand for reboiler duty calculated externally



#### 1.4.1 BLG FOR METHANOL PRODUCTION

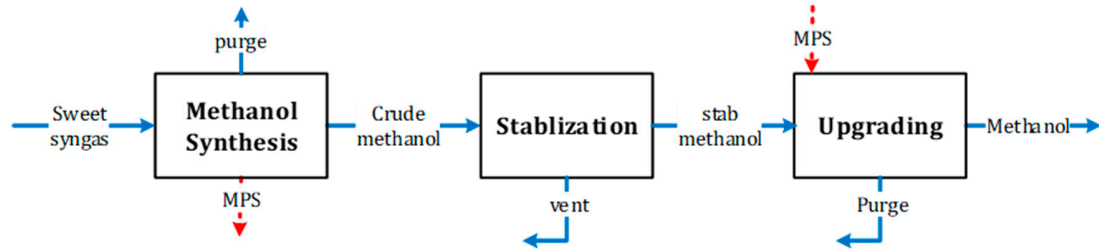


Figure S 7. Schematics of methanol synthesis from sweet syngas

Table S 7. Methanol synthesis modelling

Gas compressor	Centrifugal compressor with aftercooler to 125°C, outlet pressure 80 barg, polytropic efficiency 79%
Methanol Synthesis (MSY)	Equilibrium reactor, 233°C, 80 barg
Methanol stabilization (MST)	Multiple stage column
Methanol purification (MPU)	Distillation column, product spec 99.99 vol. % methanol

#### 1.4.2 BLG FOR FT PRODUCTION

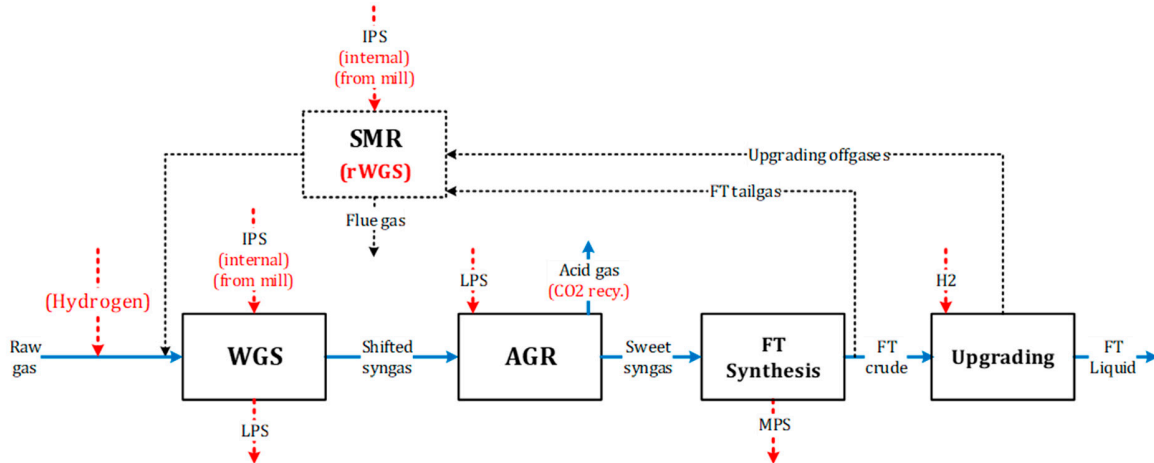


Figure S 8. Schematics of BLG for FT fuels – gas conditioning (electrification options indicated in bracket), synthesis and upgrading

Table S 8. FT fuels synthesis and upgrading modelling

FT tail gas compressor	Centrifugal compressor, outlet pressure 21 barg polytropic efficiency 79%
FT Synthesis (FTS)	Synthesis – yield reactor, maintained at 240°C, 80% internal recycle Water cooled reactor, produce 12barg saturated steam Separation – 3-phase decanter at 40°C to FTL, FT water and gases
Upgrading	Upgrading was not explicitly modelled but H <sub>2</sub> demand for upgrading is internally produced PSA modelled as component splitter
Steam reformer	Equilibrium reactor, exit temperature 950°C C <sub>1</sub> to C <sub>4</sub> assumed to be reformed

## 1.5 DUAL FLUIDIZED BED GASIFICATION (DFB)

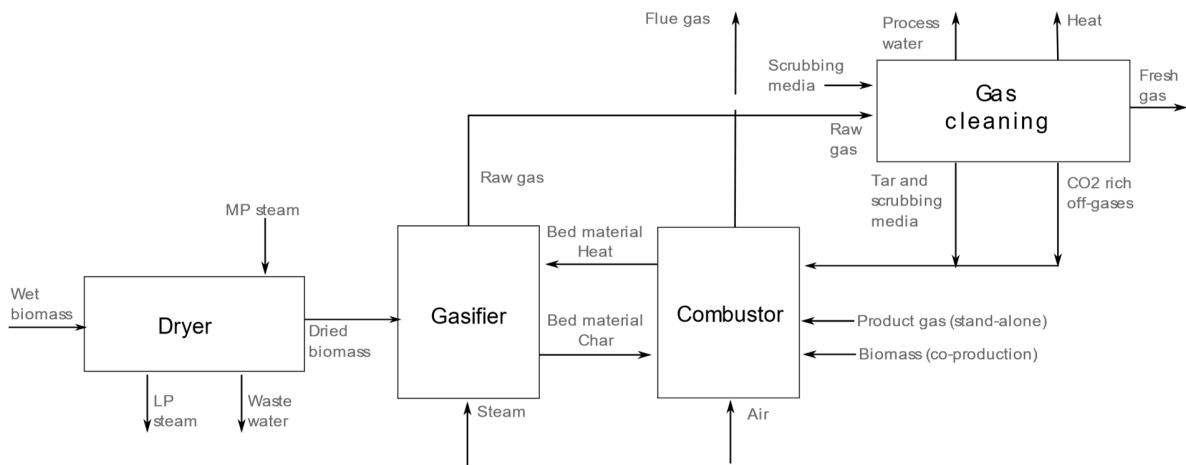


Figure S 9. Schematics of a DFB gasification system including the primary gas cleaning

## 1.6 DIRECT OXYGEN-BLOWN FLUIDIZED BED GASIFICATION (O2FB)

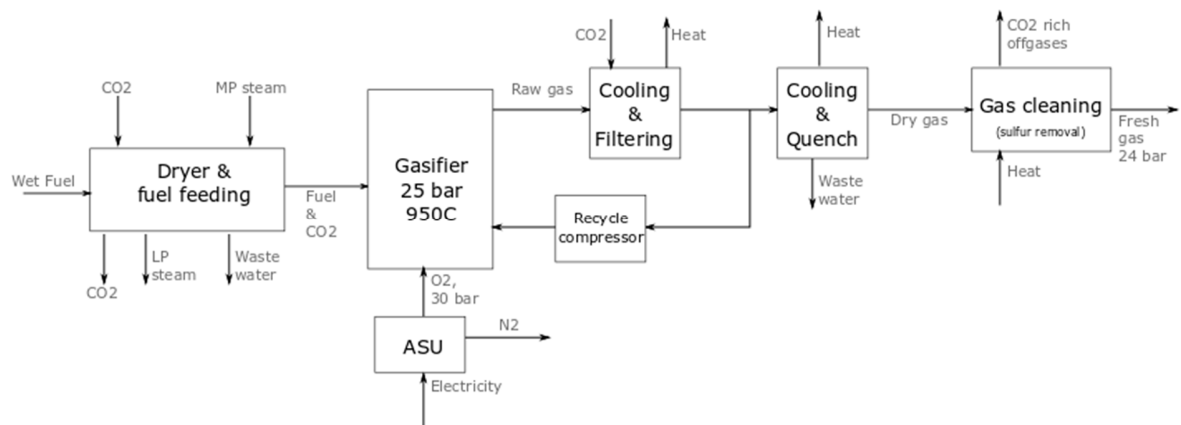


Figure S 10. Schematics of an O<sub>2</sub>FB gasifier with primary gas cleaning.

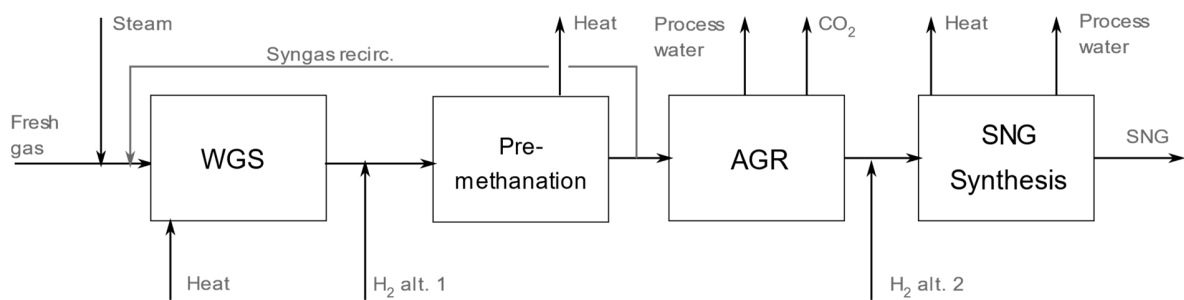


Figure S 11. Simplified process for SNG production including indications for H<sub>2</sub> addition under electrified options

Table S 9. The main assumptions used when modelling the Bio-SNG process

Main compressor (only DFB)	Outlet pressure 30 bar, 3 stage compressor. Interstage cooling down to 55 °C. Assumed isentropic efficiency: 78%
Gas cleaning (only DFB)	<p>Tar removed by scrubbing with RME, tar enriched RME sent to CHP furnace for destruction.</p> <p>MDEA Reboiler duty : 2.24 MW/kg Acid Gas [4].</p> <p>Fraction of CO<sub>2</sub> in stream co-absorbed when removing H<sub>2</sub>S: 0.1</p> <p>Assumed “complete” removal of H<sub>2</sub>S in gas, probably with help of guard beds.</p> <p>BTX separated out using activated carbon and sent to furnace. Heat losses currently neglected.</p>
WGS reactor - Low temperature catalyst	Gibbs minimization reactor. Inlet temp: 200 C.
Pre-methanator	Irreversible decomposition of higher hydrocarbons to syngas followed by equilibrium reactors. Inlet temp: given by WGS reactor and olefin content. Max allowed outlet temp 700 °C, regulated by recirculation if necessary. Subsequent cooling by raising HP steam
CO <sub>2</sub> removal – Activated MDEA Amine scrubber	Reboiler duty : 0.83 MW/kg Acid Gas [4].
SNG synthesis - 4 staged adiabatic reactors with interstage cooling	<p>Equilibrium reactors. Inlet temp 250 °C.</p> <p>The 2 last reactors are polishing steps, i.e. water is condensed out before to push the equilibrium to the necessary CH<sub>4</sub> concentration for pipeline specifications.</p> <p>Interstage cooling by HP steam raising in the first 2 reactors. LP and district heating in the polishing steps.</p>

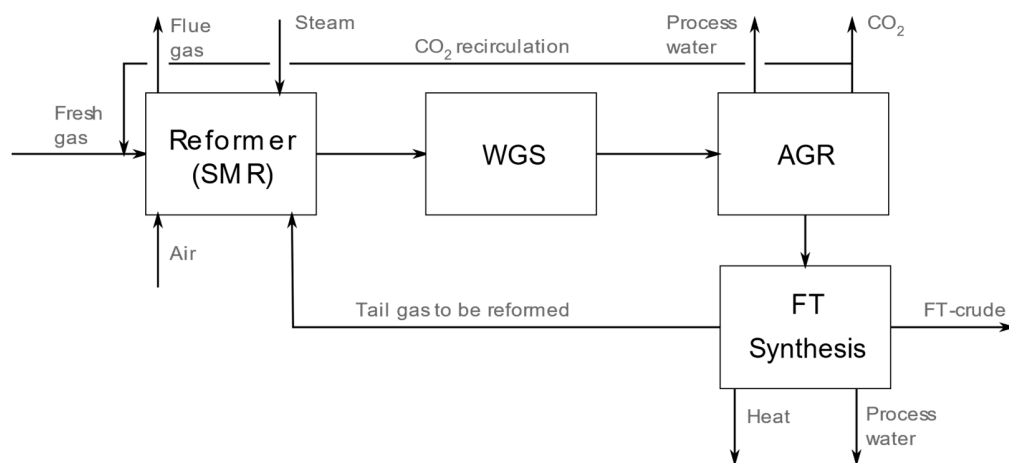


Figure S 12. Simplified process for FT production

Table S 10. The main assumptions used in the FT modelling work

Main compressor (only DFB)	Outlet pressure 30 bar, 3 stage compressor. Interstage cooling down to 55 C. Assumed isentropic efficiency: 78%
Gas cleaning (only DFB)	<p>Tar removed by scrubbing with RME, tar enriched RME sent to CHP furnace for destruction.</p> <p>MDEA Reboiler duty : 2.24 MW/kg Acid Gas [4].</p> <p>Fraction of CO<sub>2</sub> in stream co-absorbed when removing H<sub>2</sub>S: 0.1</p> <p>Assumed “complete” removal of H<sub>2</sub>S in gas, probably with help of guard beds.</p> <p>BTX separated out using activated carbon and sent to furnace. Heat losses currently neglected.</p>
Pre-reformer	<p>Adiabatic. Irreversible decomposition of higher hydrocarbons to syngas followed by Gibbs minimization.</p> <p>Inlet temp: 380 °C. Max allowed outlet temp 650 °C, regulated by CO<sub>2</sub> recirculation as thermal ballast.</p> <p>Steam ratio chosen as to give 1.8 mol H<sub>2</sub>O/mol C in hydrocarbons entering the reformer</p>
Reformer	<p>Assumed to be heated tubes with outlet at equilibrium, T<sub>out</sub> = 950°C</p> <p>Reformer heating = reaction enthalpy + heating from 800 °C to 950 °C. Uses part of fresh gas as fuel if fired, electricity in case of eSMR.</p> <p>Hydrogen added before reformer gives rise to reverse WGS and thus increased CO<sub>2</sub> utilization.</p> <p>Effluent cooled by Feed-Effluent exchange and HP steam raising</p>
WGS	Adiabatic Gibbs minimization. Inlet temp 300 °C. High temperature catalyst
CO <sub>2</sub> removal	Activated MDEA Amine scrubber. Reboiler duty : 0.83 MW/kg Acid Gas [4]. Recirculated CO <sub>2</sub> recompressed from 9 to 29 bar.
FT synthesis	<p>“Low temperature” operation at 220 °C. Alpha value at 0.9, thought to correspond to a Co-based catalyst. Production of oxygenates assumed negligible. Internal recirculation ratio of 0.7</p> <p>Partial phase separation of waxes in the reactor vessel. – flash calculation.</p> <p>Reactor cooled to maintain 220 °C by boiling water on the shell side.</p>

Phase separation	3 phase flash operating at 50 °C. Stabilizer column separating out C5 and lighter, to ensure the crude holds a vapor pressure suitable for transport. Stabilizer column not rigorously calculated, just a sharp cut.
Bleed flow	Set to 1% of the tail gas stream, to avoid N2 accumulation.
Recirculation compressor	Recompressing the FT tail gas from 22 to 28 bar to allow recirculating it back to the reformer. Assumed Isentropic efficiency: 78 %

## 2 REFERENCE E-FUELS TRACKS

Electrofuel tracks, pathways that convert electricity and CO<sub>2</sub> into fuels, are evaluated for comparison to the bio-electrofuel concepts. A total of 3 electrofuel tracks aiming at production of FT fuels, methanol, and synthetic natural gas (SNG) are considered. In the first step, syngas suitable for the synthesis of advanced biofuel is produced from electrolysis-based hydrogen (PEM) and biogenic CO<sub>2</sub> in a reverse water-gas-shift (rWGS) process operated at 750°C and 30 bar. The equilibrium of the rWGS is controlled to favor syngas composition that satisfy conditions for optimal synthesis of desired biofuel downstream.

The CO<sub>2</sub> feed is assumed to be captured from biogenic sources such as biomass-based CHP plants, pulp and paper mill, biogas, bioethanol plants. Amine process, as one of the most mature techniques for carbon capture from flue gas or other CO<sub>2</sub> containing streams, is considered for CO<sub>2</sub> capture<sup>5</sup>. The amine scrubbing technique involves a stripping column to regenerate solvent in which the CO<sub>2</sub>-rich solution flows downwards against a counter-current flow of vapor generated in the reboiler [5]. The energy demand of the reboiler makes amine scrubbing technique energy intensive. A typical MEA-based system in industrial configuration is expected to have energy consumption 3.2–4.2 MJ per kg-CO<sub>2</sub> separated, depending on the CO<sub>2</sub> concentration in feed, CO<sub>2</sub> removal rate (85%–90%), and system operating conditions [6]. Energy consumption 1MWh/tCO<sub>2</sub> (90% in form of heat and 10% in form of electricity to drive the system, excluding compression electricity) is used for the cases evaluated in this work. The heat demand of the reboiler can be satisfied by LPS (3-5 bar or 133°C -150°C) with a return temperature well above 100°C [6].

The production capacities of the electrofuel cases are selected to match available biogenic CO<sub>2</sub> sources in Sweden at scales 300 kt/y or higher [7]. 300 kt/y CO<sub>2</sub> correspond to biofuel production capacities for FT fuels (144 MW HHV), methanol (144 MW HHV) and SNG (201 MW HHV) assuming 90% annual plant availability. The mass and energy balances of the electrofuel tracks are taken from Brynolf et al. [8] for 2030 base case scenario, Table S 11.

Useful excess heat generated during synthesis is utilized to supply part of the energy demand of the amine process. Any heat deficit for the CO<sub>2</sub> removal process is assumed to be supplied from an integrated MVR heat pump that run with a COP 3. The MVR lifts temperature of water vapor generated by flashing reboiler condensate and by utilizing heat recovered from low-temperature streams available onsite, 85°C – 100°C.

Table S 11 E-fuels – relevant inputs and outputs

E-fuel tracks	Unit	eFT	eMeOH	eSNG
Input				
H <sub>2</sub>	MWh <sub>LHV</sub>	1.37	1.25	1.30
CO <sub>2</sub>	tonne	0.28	0.32	0.21
Output				
Fuel	MWh <sub>LHV</sub>	1.00	1.00	1.00



Heat	MWh <sub>th</sub>	0.20	0.10	0.20
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### 3 EQUIPMENT COST OF MAJOR PROCESS UNITS

Table S 12. Equipment cost of major process units used to derive CAPEX

Process section	Scaling parameter	Scaling exp.	Base size (S <sub>o</sub> )	Base cost (C <sub>o</sub> ) M€ <sup>a</sup>	Base year	Reference
Feedstock handling and feeding system						
Gasification feed handling system	production, MW <sub>th</sub>	0.65	100	31	2014	[9]
Belt dryer	production, MW <sub>th</sub>	0.65	100	7	2014	[9]
Gasification feeding system	production, MW <sub>th</sub>	0.60	20	6	2014	[9]
Air separation unit (ASU)	oxygen, ton/day	0.65	442	27	2017	Commercial
Liquefaction feed handling & prep	feedstock, ton <sub>wet</sub> /h	0.77	65	10	2014	[10]
Main conversion units						
Entrained flow gasifier (black liquor)	BL, tBLS/day	0.70	500	24	2017	Commercial
Dual fluidized bed gasifier	biomass feed, MW <sub>th</sub>	0.60	100	60	2020	BioShare
Fluidized bed gasifier	biomass feed, MW <sub>th</sub>	0.60	170	70	2020	BioShare
Hydrothermal liquefaction reactor	feed, ton <sub>DM</sub> /day	0.70	500	114	2014	[10]
Fast pyrolysis reactor	feed, ton <sub>DM</sub> /h	0.40	10	12	2000	[10]
Hydrogen plant	production, ton <sub>H2</sub> /day	0.70	16	31	2014	[10]
Intermediate product refining units						
Compressor	feed, kmol/h	0.70	285	4	2014 b	[9]
Conventional steam reforming (SMR)	feed, kmol/h	0.60	31 733	74	2014	[11]
Electrified steam reforming (eSMR)	feed, kmol/h	0.60	31 733	37	2014 c	
Water gas shift (WGS)	raw syngas, Nm <sup>3</sup> /h	0.65	59 000	6	2009	Commercial
Acid gas removal (amine wash)	shifted syngas, Nm <sup>3</sup> /h	0.65	15 695	2.5	2018	Commercial

Zinc bed	HHV biomass, MW <sub>th</sub>	0.65	216	2.3	2012		[12]
H <sub>2</sub> S scrubber	feed, kmol/h	0.70	285	1	2014	<sup>b</sup>	[9]
PEM electrolyser	electricity input, MW	1	1	0.8	2016	<sup>d</sup>	[13]
SOEC electrolyser	Electricity input, MW	1	1	1.2	2018		[8]
SMR/rWGS	feed, kmol/h	0.60	31 733	74	2014	<sup>e</sup>	[11]
Carbon capture (amine technology)	separated CO <sub>2</sub> , kton	1	1	0.05	2015		[8]
Mechanical vapor recompression (MVR)	heat delivered, MW <sub>th</sub>	1	1	0.5	2018		[14]
Biofuel synthesis							
Fischer Tropsch synthesis reactor	syngas feed, Nm <sup>3</sup> /h	0.75	70 630	56	2007	<sup>f</sup>	[15]
Methanol Synthesis reactor	MeOH, ton/day	0.65	465	26	2017		Commercial
Synthetic natural gas reactors	syngas feed, kmol/h	0.70	285	5.2	2014	<sup>b</sup>	[9]
Final product refining and upgrading							
HTL oil upgrading	HTL oil flow, ton/day	0.70	184	16	2014		[10]
Pyrolysis oil upgrading	Pyro oil flow, ton/day	0.70	360	57	2014		[10]
Fischer Tropsch upgrading	FT crude, ton/day	0.65	6	15	2007		[15]
Methanol upgrading	MeOH, ton/day	0.65	465	17	2017		Commercial

<sup>a</sup>Base cost other than Euro converted to Euro equivalent using average annual exchange rate of the reference year.

<sup>b</sup>Original cost reported for 20 MW biofuel product plant capacity. The scaling parameter was converted to molar flow.

<sup>c</sup>Assumed to be half of that of SMR due to the significant size reduction expected with the design of eSMR configuration [16]. Besides, no side-fired combustor is need for eSMR configuration which further reduces the capital cost compared to traditional SMR.

<sup>d</sup>PEM electrolyser cost refers to projected cost for 2030, in line with the timeline this study considers.

<sup>e</sup>Following discussions with experts in the subject, it is possible to run a steam reformer in reverse water-gas-shift (rWGS) mode by fine tuning the operating parameters (temperature, pressure, steam, catalyst) to favour desired products. The cost of rWGS is thus assumed to be the same as that of steam-methane reformer.

<sup>f</sup>FT synthesis cost recalculated to reflect the configuration used in this study. The source reported aggregated cost for FT synthesis, ATR, FT refining and recycle compressor. Cost for reformer and recycle compressor are deducted.

## 4 CARBON AND ENERGY BALANCES

Carbon and energy balances of all pathways grouped by technology track are presented here. Energy values refer to higher heating values (HHV).

### 4.1 LIGNOCELLULOSIC ETHANOL (ETOH)

Table S 13. EtOH carbon and energy balances

Scenario	EtOH_ref		EtOH_1		EtOH_2		EtOH_3	
	Carbon kg/h	Energy MW	Carbon kg/h	Energy MW	Carbon kg/h	Energy MW SOEC	Carbon kg/h	Energy MW PEM
Input								
Feedstock	13 402	137	13 402	137	13 402	137	13 402	137
Electricity		0		9		84		92
Output								
Ethanol	3 028	48	3 028	48	3 028	48	3 028	48
Biogas	658	14	658	14	2 987	61	2 987	61
Pellets	3 873	42	6 208	68	6 206	68	6 206	68
Electricity		2		0		0		0
Other								
Ferm. CO2	1 669		1 669					
AD CO2	720		720					
Flue gas	3 293		968		968		968	
District heat		4		4		5		5
Cold utility		28		12		39		47
Wastewater	160		151		213		213	

## 4.2 ALCOHOL TO JET (ATJ)

Table S 14. ATJ carbon and energy balances

Scenario	ATJ_ref		ATJ_2		ATJ_3	
	Carbon kg/h	Energy MW	Carbon kg/h	Energy MW SOEC	Carbon kg/h	Energy MW PEM
Input (background process)						
Feedstock	31 639	324	31 636	324	31 636	324
Electricity		0		197		216
Output (background process)						
Ethanol	7 136	113	7 136	113	7 136	113
Biogas/SNG export	1 552	32	7 050	145	7 050	145
Pellets	9 151	100	14 636	160	14 636	160
Electricity		6		0		0
Fermentation CO <sub>2</sub>	3 940					
AD CO <sub>2</sub>	1 699					
Flue gas	7 772		2 284		2 284	
District heat		9		3		3
Wastewater	378		519		519	
Cold utility		64		99		119
Input (ATJ)						
Ethanol	7 136	113	7 136	113	7 136	113
Biogas	938	17	0	0	0	0
Electricity		1		18		19
Output (ATJ)						
Jet	6 266	93	6 266	93	6 266	93
Diesel	820	12	820	12	820	12
Other (ATJ)						
Flue gas	987					
District heat		8		6		6
Cold utility		18		20		21
Wastewater	49		49		49	

### 4.3 HYDROTHERMAL LIQUEFACTION (HTL)

Table S 15. HTL carbon and energy balances

Scenario	HTL_ref		HTL_1		HTL_2		HTL_3		HTL_4	
	Carbon	Energy	Carbon	Energy	Carbon	Energy	Carbon	Energy	Carbon	Energy
	kg/h	MW	kg/h	MW	kg/h	MW	kg/h	MW	kg/h	MW
				SOEC		PEM		SOEC		PEM
Input										
Feedstock	7 392	94	7 392	94	7 392	94	7 392	94	7 392	94
Biogas	1 866	38	968	20	968	20				
Electricity		1		20		21		36		38
Output										
Gasoline	3 229	49	3 229	49	3 229	49	3 229	49	3 229	49
Diesel	828	11	828	11	828	11	828	11	828	11
Heavies	666	8	666	8	666	8	666	8	666	8
Electricity										
Biogas							233	5	233	5
Other										
Char	611	6	611	6	611	6	611	6	611	6
Flue gas	2 863		1 941		1 941		740		740	
AD CO2	233		233		233		233		233	
DH		9		9		9		9		9
CU &		49		50		52		42		44
losses										
Wastewater	829		852		852		852		852	

#### 4.4 FAST PYROLYSIS (FP)

Table S 16. FP carbon and energy balances

Scenario	FP_ref		FP_1		FP_2	
	Carbon kg/h	Energy MW	Carbon kg/h	Energy MW SOEC	Carbon kg/h	Energy MW PEM
Input						
Feedstock	2 354	27	2 354	27	2 354	27
Biogas	357	7	0	0	0	0
Electricity		0		7		9
Output						
Gasoline	623	10	623	10	623	10
Diesel	467	7	467	7	467	7
Heavies	236	3	236	3	236	3
Electricity		1.69		0.00		0.00
Other						
Flue gas	1 367		971		971	
District heat		1.27		1.75		1.75
CU & losses	7	13	25	14	25	15
Wastewater	12		34		34	



#### 4.5 BLACK LIQUOR GASIFICATOIN FOR METHANOL PRODUCTION (BLG-MEOH)

Table S 17. BLG-MeOH carbon and energy balances

Scenario	BLG-MeOH_ref		BLG-MeOH_1		BLG-MeOH_2		BLG-MeOH_3		BLG-MeOH_4	
	Carbon	Energy	Carbon	Energy	Carbon	Energy	Carbon	Energy	Carbon	Energy
	kg/h	MW	kg/h	MW	kg/h	MW	kg/h	MW	kg/h	MW
				SOEC		PEM		SOEC		PEM
Changes to mill BAU										
Replacement feed	25 989	291	27 064	304	27 064	304	30 165	338	30 165	338
Electricity		16		173		192		481		533
Output										
MeOH	9 542	160	16 084	270	16 084	270	27 172	456	27 172	456
Electricity		0		0		0		0		0
Secondary										
Steam to mill										
LPS		61		56		56		44		44
MPS		-2		-4		-4		-10		-10
IPS		0		0.0		0.0		0.0		0.0
District heat		4		6		6		14		14
Green liquor	3 214		3 214		3 214		3 214		3 214	
Conc. CO2	18 018		12 016		12 016		0.02		0.02	
off gases	1 398		1 615		1 615		2 490		2 490	
Cold utility		71		110		114		188		200
Wastewater	1498		740		740		794		794	
Non energy co-product										
Oxygen, kg/h		0		0		0		854		854

#### 4.6 BLACK LIQUOR GASIFICATION FOR FISCHER TROPSCH FUEL PRODUCTION (BLG-FT)

Table S 18. BLG-FT carbon and energy balance – SMR configuration

Scenario	BLG-FT_ref		BLG-FT_1		BLG-FT_2		BLG-FT_3		BLG-FT_4		BLG-FT_5	
	Carbon kg/h	Energy MW	Carbon kg/h	Energy MW	Carbon kg/h	Energy MW SOEC	Carbon kg/h	Energy MW PEM	Carbon kg/h	Energy MW SOEC	Carbon kg/h	Energy MW PEM
<b>Changes to mill BAU</b>												
Replacement feed	27 268	306	31 668	355	25 354	284	25 354	284	20 796	233	20 796	233
Electricity		6		2		126		140		514		569
<b>Output</b>												
<b>Main</b>												
FTP	7 826	120	9 057	139	13 259	204	13 259	204	24 796	381	24 796	381
Electricity		0		0		0		0		0		0
<b>Secondary</b>												
<b>Steam to mill</b>												
LPS		18		15		6		6		0		0
MPS		34		38		56		56		90		90
IPS		0		-28		1.6		1.6		2.2		2.2
District heat		29		31		22		22		38		38
Green liquor	3 201		3 201		3 201		3 201		3 201		3 201	
Conc. CO2	18 387		19 146		14 427		14 427		14		14	
Flue gas	3 654		1 497		2 024		2 024		4 944		4 944	
Cold utility		72		82		94		98		172		185
Wastewater	600		767		759		759		714		714	
<b>Other co-product</b>												
Oxygen, kg/h		0		0		0		0		958		958

Table S 19. BLG-FT carbon and energy balance – eSMR configuration

Scenario	BLG-FT_6		BLG-FT_7		BLG-FT_8		BLG-FT_9		BLG-FT_10		BLG-FT_11	
	Carbon kg/h	Energy MW	Carbon kg/h	Energy MW	Carbon kg/h	Energy MW SOEC	Carbon kg/h	Energy MW PEM	Carbon kg/h	Energy MW SOEC	Carbon kg/h	Energy MW PEM
Changes to mill BAU												
Replacement feed	27 458	308	32 851	368	27 137	304	27 137	304	28 350	318	28 350	318
Electricity		11		17		145		158		590		646
Output												
Main												
FTP	8 133	125	9 996	154	14 588	225	14 588	225	28 843	444	28 843	444
Electricity		0		0		0		0		0		0
Secondary												
Steam to mill												
LPS		16		11		2		2		0		0
MPS		35		42		60		60		75		75
IPS		0		-36		-9.7		-9.7		-29.4		-29.4
District heat		30		32		24		24		42		42
Green liquor	3 201		3 201		3 201		3 201		3 201		3 201	
Conc. CO2	18 557		19 602		14 974		14 974		1 418		1 418	
Flue gas	3 198		0		0		0		0		0	
Cold utility		73		89		103		106		209		222
Wastewater	570		870		906		906		208		208	
Other co-product												
Oxygen, kg/h		0		0		0		0		1 024		1 024

#### 4.7 FLUIDIZED BED GASIFICATION FOR SNG PRODUCTION

Table S 20. DFB-SNG marginal carbon and energy balances

Scenario	DFB-SNG_ref		DFB-SNG_1		DFB-SNG_2		DFB-SNG_3		DFB-SNG_4		DFB-SNG_5		DFB-SNG_6		DFB-SNG_7	
	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW
Biomass	3.14	115.0	3.11	113.7	2.87	105.0	2.98	109.2	2.69	98.5	2.95	107.9	2.53	92.6	2.98	109.0
RME	0.11	5.6	0.11	5.6	0.11	5.6	0.11	5.6	0.11	5.6	0.11	5.6	0.11	5.6	0.11	5.6
Electricity		9.2		10.5		37.6		33.5		86.0		75.0		174.6		151.0
SNG	1.26	90.2	1.26	90.2	1.46	105.8	1.46	105.8	1.84	132.5	1.84	132.5	2.49	181.5	2.49	181.5
DH		5.0		5.0		0.7		1.0		6.1		5.0		17.7		11.8

Table S 21. O2FB-SNG carbon and energy balances

Scenario	O <sub>2</sub> FB-SNG_ref		O <sub>2</sub> FB-SNG_1		O <sub>2</sub> FB-SNG_2		O <sub>2</sub> FB-SNG_3		O <sub>2</sub> FB-SNG_4		O <sub>2</sub> FB-SNG_5		O <sub>2</sub> FB-SNG_6	
	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW	Carbon kg/s	Energy MW
Biomass	6.69	270.0	6.69	270.0	6.69	270.0	6.69	270.0	6.69	270.0	6.69	270.0	6.69	270.0
Electricity		-0.96		23.05		19.31		287.99		248.28		391.30		338.98
SNG	2.19	160.1	2.40	174.5	2.40	174.5	4.56	329.32	4.56	329.32	5.35	387.3	5.35	387.3
DH		68.24		71.19		67.91		117.47		83.18		133.96		88.92

#### 4.8 FLUIDIZED BED GASIFICATION FOR FT FUELS PRODUCTION

Table S 22. DFB-FT marginal carbon and energy balances

Scenario	DFB-FT_ref		DFB-FT_1		DFB-FT_2		DFB-FT_3		DFB-FT_4		DFB-FT_5	
	Carbon	Energy	Carbon	Energy	Carbon	Energy	Carbon	Energy	Carbon	Energy	Carbon	Energy
	kg/s	MW	kg/s	MW	kg/s	MW	kg/s	MW	kg/s	MW	kg/s	MW
Biomass	2.70	99.0	2.76	101.1	2.85	104.4	2.85	104.4	2.85	104.2	2.85	104.2
RME	0.11	5.6	0.11	5.6	0.11	5.6	0.11	5.6	0.11	5.6	0.11	5.6
Electricity		13.53		48.58		135.40		123.82		147.75		134.68
FTP	1.07	58.96	1.54	84.69	2.30	126.90	2.30	126.90	2.41	132.57	2.41	132.57
DH		16.88		25.82		53.27		43.13		56.66		45.20

Table S 23. O2FB-FT carbon and energy balances

Scenario	O2FB-FT_ref		O2FB-FT_1		O2FB-FT_2		O2FB-FT_3	
	Carbon	Energy	Carbon	Energy	Carbon	Energy	Carbon	Energy
	kg/s	MW	kg/s	MW	kg/s	MW	kg/s	MW
Biomass (marginal)	6.69	270.0	6.69	270.0	6.69	270.0	6.69	270.0
Electricity		-6.89		59.55		343.13		305.4
FTP	1.90	104.77	2.77	152.59	5.28	286.70	5.28	286.70
DH		100.72		113.09		197.39		164.34

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