Electronic Supplementary Information

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1 Database of the 183 case studies

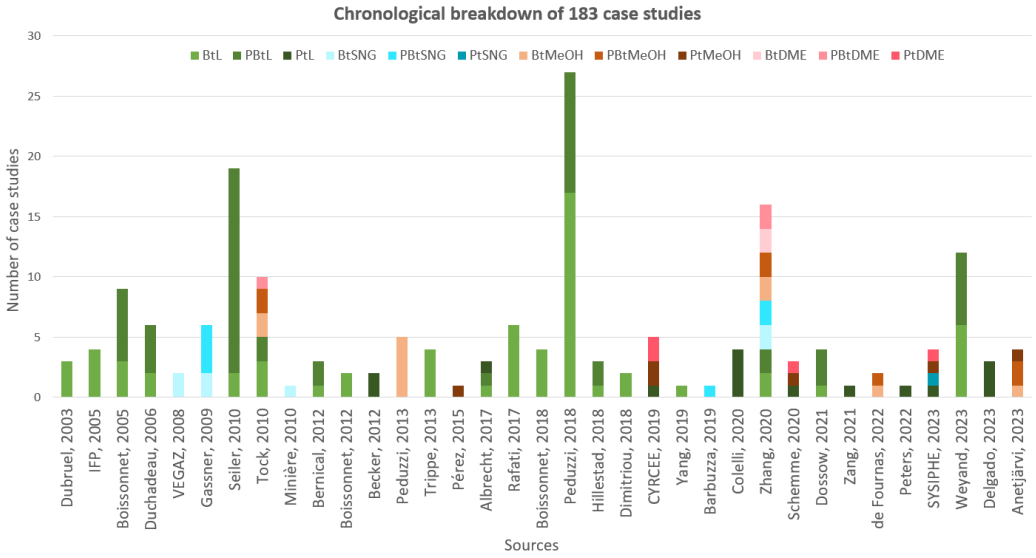
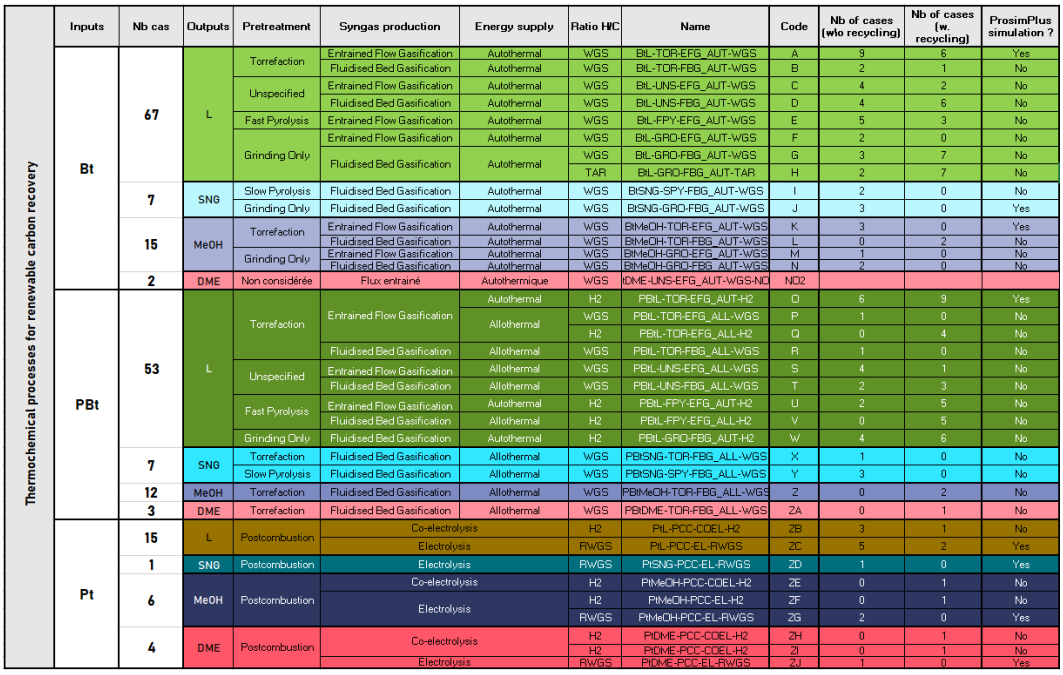


Figure 1: Chronological breakdown of 183 case studies.[22], [10], [11], [23], [**jarry˙23c˙2011**], [28], [50], [52], [39], [8], [7], [42], [53], [43], [2], [45], [12], [41], [32], [20], [37], [56], [6], [17], [58], [48], [21], [57], [25], [44], [55], [19]

2 Pathways selection



Thermochemical processes for renewable carbon recovery	Inputs	Nb cas	Outputs	Pretreatment	Syngas production	Energy supply	Ratio H/C	Name	Code	Nb of cases (w/o recycling)	Nb of cases (w. recycling)	ProsimPlus simulation ?
	Bt	67	L	Torrefaction	Entrained Flow Gasification	Autothermal	WGS	BtL-TOR-EFG_AUT-WGS	A	9	6	Yes
					Fluidised Bed Gasification	Autothermal	WGS	BtL-TOR-FBG_AUT-WGS	B	2	1	No
				Unspecified	Entrained Flow Gasification	Autothermal	WGS	BtL-UNS-EFG_AUT-WGS	C	4	2	No
					Fluidised Bed Gasification	Autothermal	WGS	BtL-UNS-FBG_AUT-WGS	D	4	6	No
				Fast Pyrolysis	Entrained Flow Gasification	Autothermal	WGS	BtL-FPY-EFG_AUT-WGS	E	5	3	No
				Grinding Only	Entrained Flow Gasification	Autothermal	WGS	BtL-GRO-EFG_AUT-WGS	F	2	0	No
					Fluidised Bed Gasification	Autothermal	WGS	BtL-GRO-FBG_AUT-WGS	G	3	7	No
							TAR	BtL-GRO-FBG_AUT-TAR	H	2	7	No
		7	SNG	Slow Pyrolysis	Fluidised Bed Gasification	Autothermal	WGS	BtSNG-SPY-FBG_AUT-WGS	I	2	0	No
				Grinding Only	Fluidised Bed Gasification	Autothermal	WGS	BtSNG-GRO-FBG_AUT-WGS	J	3	0	Yes
		15	MeOH	Torrefaction	Entrained Flow Gasification	Autothermal	WGS	BtMeOH-TOR-EFG_AUT-WGS	K	3	0	Yes
					Fluidised Bed Gasification	Autothermal	WGS	BtMeOH-TOR-FBG_AUT-WGS	L	0	2	No
				Grinding Only	Entrained Flow Gasification	Autothermal	WGS	BtMeOH-GRO-EFG_AUT-WGS	M	1	0	No
					Fluidised Bed Gasification	Autothermal	WGS	BtMeOH-GRO-FBG_AUT-WGS	N	2	0	No
		2	DME	Non considérée	Flux entrainé	Autothermique	WGS	tDME-UNS-EFG_AUT-WGS-NO	NO2			
	PBt	53	L	Torrefaction	Entrained Flow Gasification	Autothermal	H2	PBtL-TOR-EFG_AUT-H2	O	6	9	Yes
						Allothermal	WGS	PBtL-TOR-EFG_ALL-WGS	P	1	0	No
							H2	PBtL-TOR-EFG_ALL-H2	Q	0	4	No
					Fluidised Bed Gasification	Allothermal	WGS	PBtL-TOR-FBG_ALL-WGS	R	1	0	No
				Unspecified	Entrained Flow Gasification	Allothermal	WGS	PBtL-UNS-EFG_ALL-WGS	S	4	1	No
					Fluidised Bed Gasification	Allothermal	WGS	PBtL-UNS-FBG_ALL-WGS	T	2	3	No
				Fast Pyrolysis	Entrained Flow Gasification	Autothermal	H2	PBtL-FPY-EFG_AUT-H2	U	2	5	No
					Fluidised Bed Gasification	Allothermal	H2	PBtL-FPY-EFG_ALL-H2	V	0	5	No
				Grinding Only	Fluidised Bed Gasification	Autothermal	H2	PBtL-GRO-FBG_AUT-H2	W	4	6	No
		7	SNG	Torrefaction	Fluidised Bed Gasification	Allothermal	WGS	PBtSNG-TOR-FBG_ALL-WGS	X	1	0	No
				Slow Pyrolysis	Fluidised Bed Gasification	Allothermal	WGS	PBtSNG-SPY-FBG_ALL-WGS	Y	3	0	No
		12	MeOH	Torrefaction	Fluidised Bed Gasification	Allothermal	WGS	PBtMeOH-TOR-FBG_ALL-WGS	Z	0	2	No
		3	DME	Torrefaction	Fluidised Bed Gasification	Allothermal	WGS	PBtDME-TOR-FBG_ALL-WGS	ZA	0	1	No
	Pt	15	L	Postcombustion	Co-electrolysis		H2	PtL-PCC-COEL-H2	ZB	3	1	No
					Electrolysis		RWGS	PtL-PCC-EL-RWGS	ZC	5	2	Yes
		1	SNG	Postcombustion	Electrolysis		RWGS	PtSNG-PCC-EL-RWGS	ZD	1	0	Yes
		6	MeOH	Postcombustion	Co-electrolysis		H2	PtMeOH-PCC-COEL-H2	ZE	0	1	No
					Electrolysis		H2	PtMeOH-PCC-EL-H2	ZF	0	1	No
							RWGS	PtMeOH-PCC-EL-RWGS	ZG	2	0	Yes
		4	DME	Postcombustion	Co-electrolysis		H2	PtDME-PCC-COEL-H2	ZH	0	1	No
							H2	PtDME-PCC-COEL-H2	ZI	0	1	No
					Electrolysis		RWGS	PtDME-PCC-EL-RWGS	ZJ	1	0	Yes

Figure 2: Specifications of the 37 identified thermochemical conversion processes.

3 Modelling assumptions

Biomass modelling

Biomass is not an existing component in process simulation databases. It has thus been modeled using three existing components, introduced in selected proportions to obtain a mixture with an average atomic composition close to that of lignocellulosic biomass $C_6H_9O_4$. By measuring the LHV of the mix (18.08 MJ.kg^{-1}), we define standard enthalpies of formation of the 3 pseudo-components.

Resource processing

Torrefaction is a pre-treatment phase preceding the transportation and gasification of lignocellulosic biomass. It allows for an increase in the H_2 and CO content of syngas by decreasing the O/C ratio of the initial resource [18] and improves biomass grindability [54]. The modeling assumptions for torrefaction are based on two previous internal studies. It involves a simple reactor at 250°C for the torrefaction reaction and a simple reactor at 850°C for the combustion of the fumes produced by the torrefaction. Grinding is a size reduction step required for entrained-flow gasifier technologies (see 3). This

stage is not modeled in the simulations but is included in the economic analysis. E-fuel processes include a post-combustion carbon capture step. It contains a CO_2 absorption phase using an amine solvent at 60°C and a desorption phase to separate the CO_2 and recycle the amine solvent at 110°C. Carbon capture energy requirements are concentrated in the desorption column reboiler [38]. This carbon capture model is inspired by an open-source process example developed by Prosim [15].

Syngas production

Two main biomass gasification technologies coexist in the literature. Fluidized bed gasifiers operate between 800°C and 900°C with a physical separation of the gasification and combustion reactions. In particular, the fluidized bed allows the solid residue formed during gasification to flow into the combustion zone. The separation of the two chambers also enables the separation of the two gas streams produced: syngas and flue gas. This type of technology co-produces a large quantity of methane and is therefore favored for SNG production [39]. Entrained-flow reactors are better suited for high-power industrial plants ($> 50MWth$). Biomass is injected along with gaseous reactants in a circulating furnace, resulting in partial combustion of the resource. Usually operating at high temperatures (1200 - 1400 °C) to facilitate tar reduction, the entrained-flow reactor is modeled by a balanced reactor minimizing Gibbs free energy, at 1400 °C. Hydrogen production is used in e-fuel and e-biofuel processes as energy carriers for fuel production. A solid oxide electrolysis cells (SOEC) module developed in a previous study [37] is chosen to carry out the simulation. Operating at a high temperature (800°C) means that part of the electrical input required can be replaced by a thermal input, improving the process's energy integration potential [8]. Simulations of biofuel and e-biofuel processes do not require any external heat input, as they produce excess heat power at both low and high temperatures. On the other hand, e-fuel processes require a high-temperature heat input for electrolysis. In the case of the PtL process, this input is provided by the combustion of tail gas at the outlet of the Fischer-Tropsch reactor. For the PtSNG and PtMeOH processes, the heat input is provided by the combustion of the respective products, mainly methane and methanol.

H/C ratio adjustment

The H/C ratio is defined as the H_2 mole fraction over the CO mole fraction. Fuel synthesis requires a specific H/C ratio for optimal performance, between 1.6 and 2.1 depending on the synthesis. For biofuel processes BtL, BtMeOH, BtSNG, H_2 proportion in the syngas needs to be increased. The desired ratio is readjusted with the water-gas-shift (WGS) reaction:

$$CO + H_2O \rightleftharpoons H_2 + CO_2 \quad (1)$$

With steam addition and CO molecules, this exothermic reaction ($\Delta H_r^0 = -42$ kJ.mol^{-1}) produces extra H_2 and CO_2, resulting in a H/C ratio increasing along with CO_2 losses in the process. The WGS reactor is modeled by a simple reactor operating at 385°C. In the

PtL process, the desired ratio is achieved by shifting the WGS reaction to transform CO_2 into CO, consuming part of the hydrogen supply. This "reverse water-gas shift" reaction requires high pressure and temperature conditions: 25 bar, 600°C [46]. For hybrid processes PBtL, PBtSNG, and PBtMeOH, the quantity of H_2 produced by electrolysis is adjusted to reach the optimal syngas ratio for the fuel synthesis, avoiding CO_2 losses from the WGS reaction.

Fuel synthesis

The liquid hydrocarbon production uses a Fischer-Tropsch reactor, which first converts syngas into C_1 to C_{30} hydrocarbons. It operates at 220°C in the presence of a cobalt or iron catalyst. The Anderson-Schulz-Flory (ASF) distribution is chosen to describe hydrocarbon production in the Fischer-Tropsch reactor during the polymerization process [36]. The chain growth probability α is set to 0.9 to maximize the production of C_5-C_{20} carbon chains (naphtha, kerosene, gasoline, diesel). A light gas recycling loop and heavy waxes hydrocracking reactor are modeled to convert C_1-C_4 and C_{20+} hydrocarbon chains, increasing production yields [57]. SNG production is carried out with a methanation unit including an olefine reformer and a methanation reactor, both modeled by a constant temperature reactor at 385°C. The methane synthesis is exothermic, which allows heat to be recovered in turbines and provides electrical power for auxiliaries consumption. It involves two main reactions converting CO_2 (Sabatier reaction) or CO (methanation). The methanation unit simulation was previously developed for two internal studies. Finally, methanol synthesis is simulated with a methanolization unit, with a reactor minimizing Gibbs free energy and liquid-vapor separators. The model derives from an open-source process example developed by Prosim and an internal project [37].

Software, thermodynamic model and reactions

Simulations are carried out using a commercial process simulation software – ProSimPlus® – to allow physical modeling as well as mass and energy balances; modeling is based on standard elementary and user modules and supported by various CEA (French Atomic Energy and Alternative Energies Commission) previous works. The main operating conditions and reactions in each modeled unit are available in Table 1.

4 Performance criteria

Carbon conversion

Biomass and CO_2 from concentrated sources can be considered as carbon resources with limited potential [12]. These limits initially raise the question of how best to allocate these resources prioritizing uses. This issue is especially true for biomass resources, with potential conflicts with the agriculture and forestry sectors [49]. Secondly, it is crucial to

Table 1: Main reactions and operating conditions

Step	**Reaction(s)**	$P(\mathbf{bar}), T(^{\circ}\mathbf{C})$
Torrefaction	See section 5.1	1.03, 280
Comb. Biomass 1	$C_6H_{10}O_4 + 6.5\ O_2 \rightarrow 6\ CO_2 + 5\ H_2O$	1.05, 850
Comb. Biomass 2	$C_6H_6O_3 + 6\ O_2 \rightarrow 6\ CO_2 + 3\ H_2O$	1.05, 850
Comb. Biomass 3	$C_6H_{12}O_6 + 6\ O_2 \rightarrow 6\ CO_2 + 6\ H_2O$	1.05, 850
Comb. Methane	$CH_4 + 2\ O_2 \rightarrow CO_2 + 2\ H_2O$	1.05, 850
Comb. Acetic acid	$CH_3COOH + 2\ O_2 \rightarrow 2\ CO_2 + 2\ H_2O$	1.05, 850
Gasification EFG	Gibbs free energy minimization	35, 1400
Gasification FICB	See section 5.2	1.5, 800
Water-gas shift	$CO + H_2O \rightarrow CO_2 + H_2$	35, 250
Reverse water-gas shift	$CO_2 + H_2 \rightarrow CO + H_2O$	25, 600
Fischer-Tropsch	$nCO + (2n+1)\ H_2 \rightarrow n\ H_2O + C_nH_{2n+2}$ (ASF)	35, 220
Hydrocracking	See section 5.3	50, 50
Methanolization	Gibbs free energy minimization	110, 150
Methanation CO	$CO + 3\ H_2 \rightarrow CH_4 + H_2O$	1, 385
Sabatier reaction	$CO_2 + 4\ H_2 \rightarrow CH_4 + 2\ H_2O$	1, 385

value limited carbon in the most efficient way possible. This efficiency based on molar flow rates is known as the carbon conversion ratio and is defined as follows:

$$\eta_C = \frac{\dot{m}_{fuels}.M(feedstocks).\overline{N}_{C-fuels}}{\dot{m}_{feedstocks}.M(fuels).\overline{N}_{C-feedstocks}} \tag{2}$$

where $\dot{m}_{fuels}$ denotes the mass flow rate of the output products (liquid hydrocarbons, synthetic natural gas, methanol) in kg.s^{-1}, $\dot{m}_{feedstocks}$ is the mass flow rate of the carbon feedstock, alternatively biomass or carbon dioxide, $M(fuels)$ and $M(feedstocks)$ are the molar mass of the different fuels and feedstocks in g.mol^{-1}, $\overline{N}_{C-fuels}$ and $\overline{N}_{C-feedstocks}$ are the average number of carbon atoms per molecule of fuel or feedstock. Other indicators are frequently used in the literature, such as material efficiency or chemical energy efficiency [50][41][52][17][20]. They derive directly from the first definition.

Energy efficiency

Energy yield is the performance indicator used to quantify the energy recovery of the resource in the product. It may or may not include auxiliary energy consumption required for process operation. Three definitions coexist in the literature. The internal energy yield η_{int} is expressed as follows:

$$\eta_{int} = \frac{\dot{m}_{fuels}.LHV_{fuels}}{\dot{m}_{biomass}.LHV_{biomass} + P_{electrolysis}} \tag{3}$$

where $\dot{m}_{fuels}$ denotes the mass flow rate of the output products (liquid hydrocarbons, synthetic natural gas, methanol) in kg.s^{-1}, LHV_{fuels} is the Lower Heating Value (LHV)

of the output products in MJ.kg^{-1}, $\dot{m}_{biomass}$ is the dry biomass input mass flow rate in kg.s^{-1}, $LHV_{biomass}$ is the LHV of the dry biomass, and $P_{electrolysis}$ is the electrical power to be supplied for hydrogen production by electrolysis in MW. In biofuel processes $P_{electrolysis} = 0$ whereas in e-fuel processes $\dot{m}_{biomass} = 0$. The overall energy yield $\eta_{ov.}$ is defined by the following formula:

$$\eta_{ov.} = \frac{\dot{m}_{fuels}.LHV_{fuels} + P_{prod-elec}}{\dot{m}_{biomass}.LHV_{biomass} + P_{electrolysis} + P_{aux-elec} + P_{aux-therm}} \quad (4)$$

This efficiency integrates the electrical power recovered from waste heat by turbines $P_{prod-elec}$ and consumed by auxiliary units (compressors, pumps, reboiler) $P_{aux-elec}, P_{aux-therm}$. As most frequently encountered in the literature, we used this efficiency indicator to assess technical performance in this study. After energy integration, potential surplus waste heat is not considered in the calculation of energy yield.

Among the 184 cases reviewed, 174 apply this definition, while 96 use a "primary energy efficiency" that applies a correction factor to electricity (typically 2.6–3). Although such factors reflect the conversion losses of thermal power generation, their applicability depends on system boundaries and electricity generation mixes, and may introduce a bias against electrified pathways. In our context, the indicator is process-based and focuses on the energy content of the produced fuels, which is particularly relevant for hard-to-electrify sectors such as aviation. This metric should therefore be interpreted as one criterion among others in a multi-criteria framework (energy, carbon, economics, environment), and not as a unique measure of energy usefulness. Exergy-based approaches could provide complementary insights, but fall beyond the scope of the present analysis.

However, to ensure comparability of processes on a primary energy basis, we included a calculation of energy efficiencies assuming four different origins of electricity production: the average European electricity mix, nuclear power, a renewable mix (33% photovoltaic, 33% wind, and 33% hydropower), and combined heat and power (CHP). Accordingly, the electricity consumption of each process was converted using a distinct primary energy factor (PEF) depending on the source of electricity. The selected PEF values and corresponding references are presented in Table 2.

Table 2: Primary energy factors (PEF) used for different electricity sources.

Electricity source	**PEF**	**Ref.**
Average EU electricity mix	2.0	[5]
Nuclear power	3.0	[26]
Renewable mix (PV + wind + hydropower)	1.0	[9]
Combined heat and power (CHP)	1.25	[1]

5 Reactions

5.1 Torrefaction

$$55.8\,C_6H_{10}O_4 + 30.1\,C_6H_6O_3 + 14.1\,C_6H_{12}O_6 \rightarrow 73.22\,C_6H_8O_4 + 20.34\,C_6H_{10}O_3 + 31.36\,H_2O + 7.07\,CO_2$$
$$+ 2.69\,CH_3COH + 2.15\,CH_3COOH + 1.73\,CO + 1.5\,CH_2O_2 + 0.38\,C_3H_6O_3 + 0.27\,C_5H_4O_2 + 0.17\,C_3H_6O_2 + 0.14\,O_2$$

5.2 FICB Gasification

$$0.483\,C_6H_6O_3 + 0.400\,C_6H_{12}O_6 + 0.117\,C_6H_{12}O_3 + 1.099\,H_2O \rightarrow$$
$$0.130\,C_{14}H_8O_2 + 0.756\,CH_4 + 3.261\,H_2 + 1.880\,CO + 1.484\,CO_2 + 0.037\,C_2H_6 + 0.075\,C_2H_4 + 0.037\,C_2H_2 + 0.004\,C_{10}H_8$$

5.3 Hydrocracking

$$H_2 + 0.379\,C_{36}H_{74} + 0.285\,C_{32}H_{66} + 0.259\,C_{24}H_{50} + 0.189\,C_{28}H_{58} \rightarrow$$
$$0.126\,CH_4 + 0.101\,C_2H_6 + 0.091\,C_3H_8 + 0.121\,C_4H_{10} + 0.168\,C_5H_{12}$$
$$+ 0.222\,C_6H_{14} + 0.252\,C_7H_{16} + 0.274\,C_8H_{18} + 0.299\,C_9H_{20} + 0.053\,C_{10}H_{22} + 0.965$$
$$C_{12}H_{24} + 0.5222\,C_{15}H_{32} + 0.164\,C_{18}H_{38} + 0.062\,C_{20}H_{42}$$

6 Economic analysis

6.1 Indicators

Investment costs

Second-generation-type biofuel processes, e-biofuel, and e-fuel processes require heavy transformation infrastructures. The capital intensity C_{int} has a significant impact on the expected production cost. This crucial indicator for potential investors is expressed as follows:

$$C_{int} = \frac{C_{dep}}{Capacity} \tag{5}$$

with C_{dep} the depreciable capital cost including the cost of equipment, installation, and working capital in €$_{2023}$ (section 6.2). Installed capacity corresponds to the product output power on LHV basis in kW.

Net production costs

The Net Production Cost (NPC) is useful to compare the estimated competitiveness of the different sectors, relative to market prices. It is calculated as follows:

$$NPC = \frac{C^{fixed} + C^{variable}}{Production_n} \tag{6}$$

Fixed costs (C^{fixed}) include the depreciation of invested capital, while variable costs ($C^{variable}$) include the cost of inputs (biomass, electricity). The annual production volume of year n $Production_n$ is calculated from material balances of process simulations. The calculation of annualized operating costs is further detailed in section 6.2.

6.2 Economic modeling

The calculation of economic indicators is conducted following the typical economic methodology available in figure 7) [16].

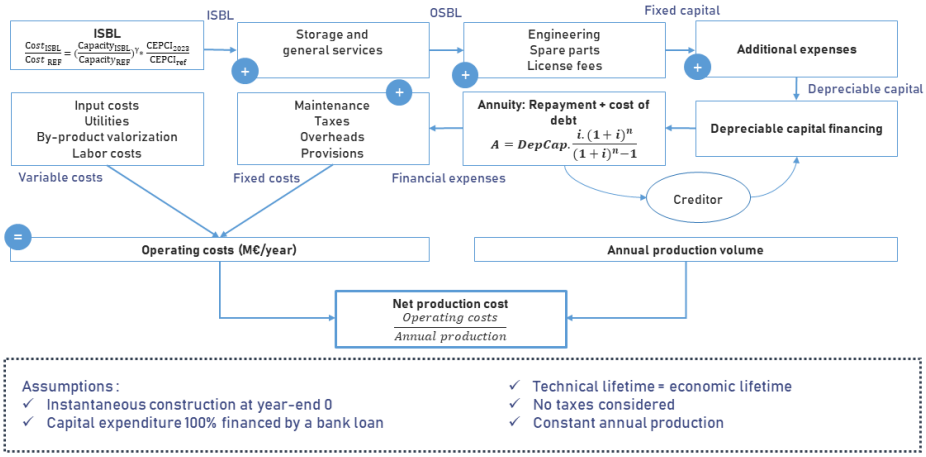


Figure 3: Net production cost calculation methodology

Equipment costs

The cost of industrial equipment is estimated with a reference cost and a scale factor γ allowing extrapolation relative to the capacity of the installed equipment. Costs were updated using values from the Chemical Engineering Plant Cost Index (CEPCI). In the study, we gathered several reference costs for the same industrial equipment. Values provided by different studies can often differ significantly, depending on the chosen assumptions. An arbitration was conducted to select a reference value for each technological component of our processes. The scale factor is specific to each equipment and is usually specified in the source study. In the absence of details, we applied a standard value of 0.7 for the scale factor. The list of equipment required in the simulated processes and their associated reference costs are specified in Table 3.

The "Inside Battery Limit" (ISBL) cost of each piece of equipment is defined as

follows :

$$\frac{\text{Cost}_{\text{ISBL}}}{\text{Cost}_{\text{Ref.}}} = \left(\frac{\text{Capacity}_{\text{ISBL}}}{\text{Capacity}_{\text{Ref.}}}\right)^{\gamma} \times \frac{\text{CEPCI}_{2023}}{\text{CEPCI}_{\text{Ref.}}} \tag{7}$$

Due to a lack of data, extrapolations in this study may exceed the recommended duration of 5 years for cost updating. This recommendation does not seem to be particularly respected in the literature.

Investment costs

In addition to ISBL costs, the fixed capital cost includes the cost of utilities and control, piping, storage, buildings and infrastructure, engineering, risk provisions, and licenses. According to [29] the factor between the ISBL cost and fixed capital is typically ranging from 1.8 to 2.5 in the associated literature. The estimation of each expense item is typically expressed as a percentage of the ISBL cost. By cross-comparison of assumptions made by studies in the literature with those chosen in [16], typical values are selected for this study and presented in Table 8.

Under these assumptions, the installation factor between the ISBL equipment cost and the fixed capital reaches 2.25. The depreciable capital is obtained by adding startup costs and loan interests.

Industrial-scale Fischer-Tropsch units generally have input capacities between 200 and 500 MW, corresponding to dry biomass inputs between 40 and 100 t/h for biofuel processes [24]. On average, SNG synthesis units are smaller with input capacities between 20 and 100 MW [35]. Finally, for methanol synthesis, we find rather low values of around 20 MW for BtMeOH and PBtMeOH processes [42] and rather high values of around 500 MW for PtMeOH processes [43]. These values correspond to capacities for industrial units and are not based on real operational units, as the industrial deployment of this type of technology is still in its infancy [14]. The sizing of each installation is detailed in the Table 6.

Fixed costs

As most frequently encountered in the literature, a net production cost (NPC) calculation is chosen over the Levelized Cost of Production (LCOP) method [30]. Constant annual operating costs approximation is valid under specific assumptions :

- The construction of the facilities is instantaneous, at the end of year 0.
- Depreciable capital is 100% financed by debt, with a chosen fixed interest rate.
- The loan repayment period is equal to the depreciation period of the capital and the technical lifespan of the facilities.
- No taxes are applied.

- Annual fuel production is assumed to be constant over the entire lifespan of the facilities.

Under these assumptions the debt is repaid with interest in the form of annuities, calculated as follows:

$$A = C_{dep} \cdot \frac{i \cdot (1+i)^n}{(1+i)^n - 1} \tag{8}$$

A is the annuity representing the financial burden, C_{dep} is the value of the depreciable capital of the project in million euros (M€), i is the bank loan interest rate, and n is the lifespan and repayment period of the installation in years. Other fixed costs are taken into account: maintenance, property taxes, insurance, and overhead. The values chosen for these expenses are based on those most commonly found in the literature and are specified in Table 4. They are expressed as a fraction of the ISBL (+ Utilities) cost or fixed capital. Finally, the cost of labor is estimated by considering a need for k operators (where k varies depending on the size of the unit), remunerated at 60 k€per year, and present on-site for 8 hours per day in shifts. To take into account rest times, we applied 4.56 multiplication factor [27].

Variable costs

In the context of our study, the primary inputs include dry biomass, electricity, natural gas, and water. Paraffin wax is also used in the BtSNG and PBtSNG processes during the fluidized bed reactor gasification stage. The cost of catalysts is disregarded. Dry biomass is traded in the form of woodchips. The woodchip purchase price reflects the costs of cutting, drying, grinding, and transporting the resource. It is set in the study as a frequently encountered literature value [2][41][3]. The prices of electricity and natural gas are average market prices (EPEX spot for electricity, PEG for natural gas) for the year 2023. The paraffin wax price is taken from the Paraffin price index, in Europe.

7 Reference costs and dimensions of equipment

Table 3: Reference costs and dimensions of the main transformation units.

Unit	**Cost (M€)**	**Size**	**Unit**	γ	**Year**	**CEPCI**	**Source**
Torrefaction	13,21	100	$kt.year^{-1}$	0,7	2011	585,7	[25]
Grinding	0,75	200	MW_{th} LHV	0,6	2002	395,6	[45]
Drying	16,69	200	MW_{th} LHV	0,8	2002	395,6	[45]
Carbon capture	58,01	2709	kmol $CO_2.h^{-1}$	0,7	2012	708,8	[4]
EF reactor	54,59	432	MW_{th}	0,7	2012	584,6	[40]
FICB reactor	35,43	68,8	$t.h^{-1}$	0,7	2001	394,3	[31]
Quench	0,33	3,743	$m^3.s^{-1}$	0,67	2012	584,6	[40]
ASU	29,11	576	$t.day^{-1}$	0,75	2012	584,6	[40]
WGS	13,39	8819	$kmol.h^{-1}$	0,65	2002	395,6	[25]
RWGS	37,47	4000	$kmol.h^{-1}$	0,7	2019	607,5	[57]
SOEC	0,48	1	MW_{el}	1	2019	607,5	[34]
AGR	4,54	542	kmol $CO_2.h^{-1}$	0,75	2012	584,6	[40]
FT reactor	11,21	0,9025	$kmol.s^{-1}$	0,67	2014	576,1	[40]
SNG synthesis	5,23	175	MW_{out} HHV	0,75	2012	584,6	[33]
MeOH synthesis	76,05	10,81	$kmol.s^{-1}$	0,7	2002	395,6	[33]
Hydrocracking	12,00	19,7	$t.h^{-1}$	0,7	2010	550,8	[51]
Compressor	6,31	10	MW_{el}	0,67	2007	525,4	[25]
Exchanger	18,10	355	MW_{th}	0,7	2007	525,4	[25]
MeOH purification	1,60	4,66	$kg.s^{-1}$	0,291	2002	395,6	[33]

8 Additional costs

Expense	**Cost**
Utilities and control	0.2 ISBL
Piping	0.15 ISBL
Storage	0.25 ISBL
Building and infrastructure	0.2 ISBL
Engineering	0.2 ISBL
Contingencies	0.25 ISBL

Table 4: Expense breakdown and corresponding costs.

9 GHG analysis assumptions

Feedstock	**Emission factor ($kgCO_{2,eq}$/MWh)**	**Reference**
Biomass (< 500 km)	26.4	RED III
Electricity (2022, France)	54.4	RTE
Electricity (2050, France)	13.0	RTE [47]
Electricity (2023, EU average)	326	RTE
Electricity (nuclear)	6.0	UNECE, 2022
Electricity (33% PV, 33% wind, 33% hydro)	18	UNECE, 2022
Electricity (CHP)	322	[13]
Natural gas	197.8	Chiffres clés du climat, 2023
Waxes	327.6	ADEME, Base Carbone

Table 5: Reference and alternative emission factors for the different contributing feedstocks.

10 Simulation input sizing

Table 6: Biomass and power inputs of the 9 processes.

Process	**Biomass input (MW_{th})**	**Electrolysis input (MW_{el})**
BtL	436	0
PBtL	436	268
PtL	0	566
BtSNG	67	0
PBtSNG	67	57,8
PtSNG	0	57,8
BtMeOH	436	0
PBtMeOH	436	244
PtMeOH	0	501

11 Sensitivity on energy efficiency

Table 7: Primary energy efficiency of each process under different electricity supply assumptions and their primary energy factor (PEF).

Process	**Primary energy efficiency**			
	EU mix (PEF 2.0)	**Nuclear** (PEF 3.0)	**Renewable mix** (PEF 1.0)	**CHP** (PEF 1.25)
BtL	47.1%	45.9%	48.3%	48.0%
PBtL	41.5%	32.6%	57.3%	52.4%
PtL	26.1%	17.9%	48.5%	39.9%
BtSNG	46.5%	45.7%	47.4%	47.2%
PBtSNG	45.9%	36.3%	64.0%	58.1%
PtSNG	35.6%	25.5%	61.2%	51.6%
BtMeOH	52.5%	50.5%	54.7%	54.2%
PBtMeOH	45.9%	35.9%	63.5%	58.0%
PtMeOH	31.8%	22.4%	54.7%	46.4%

Note: PEF stands for Primary Energy Factor, which converts final electricity into its primary energy equivalent.

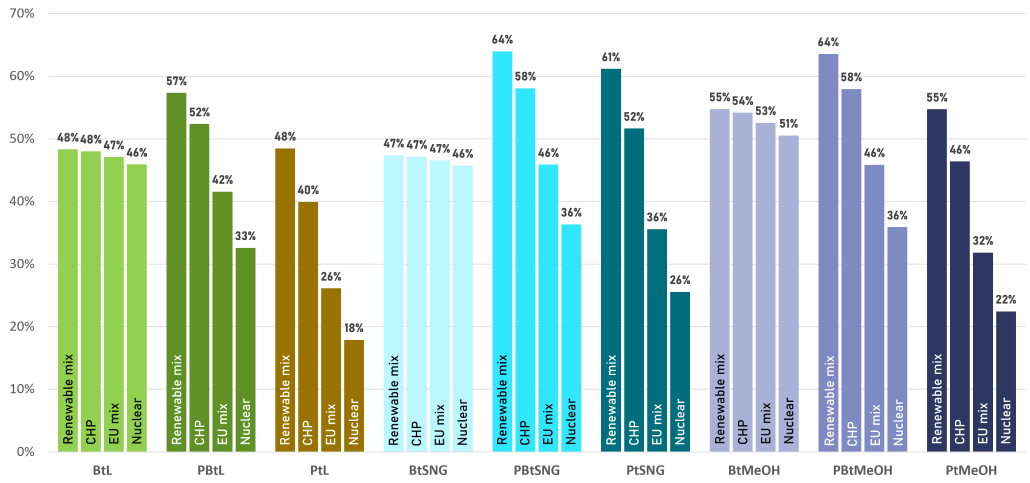
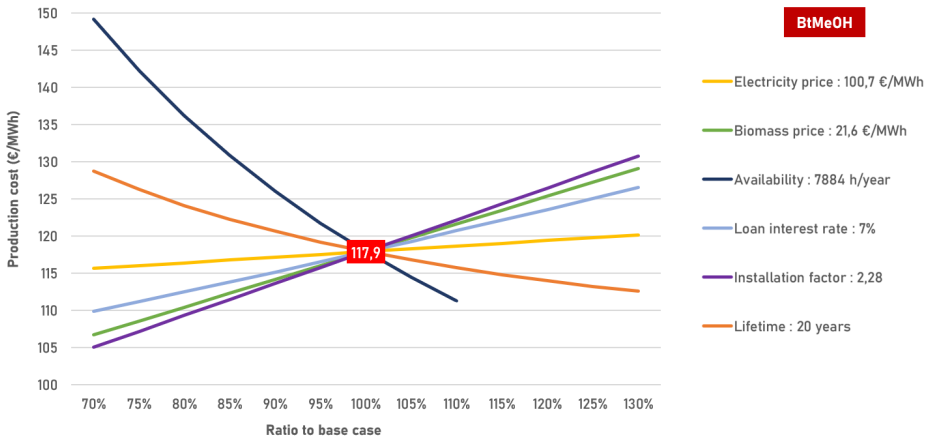
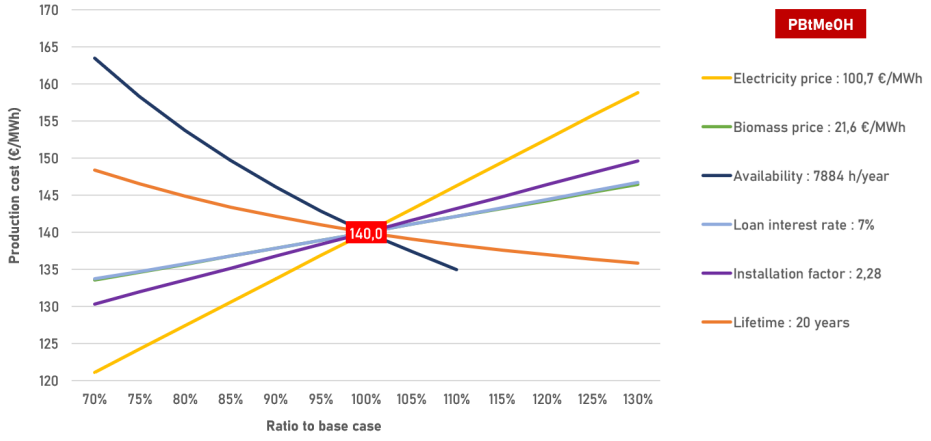


Figure 4: Evolution of the primary energy efficiency of each process under different electricity supply assumptions.

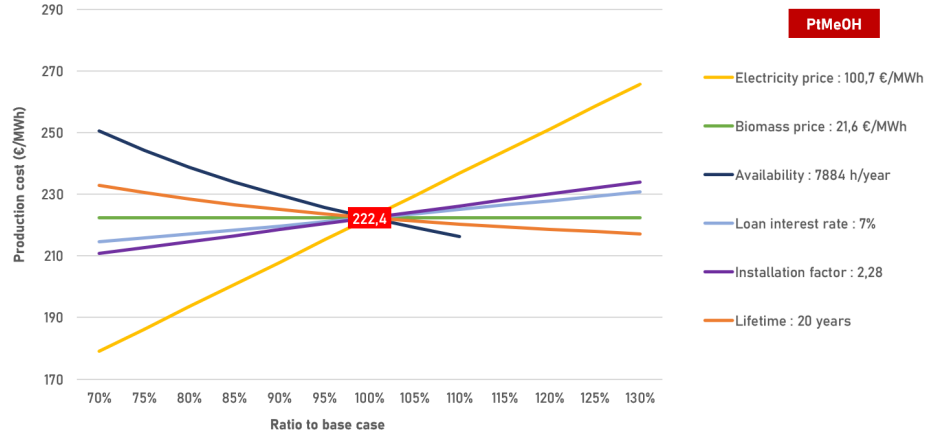
12 Production costs sensitivity



(a)

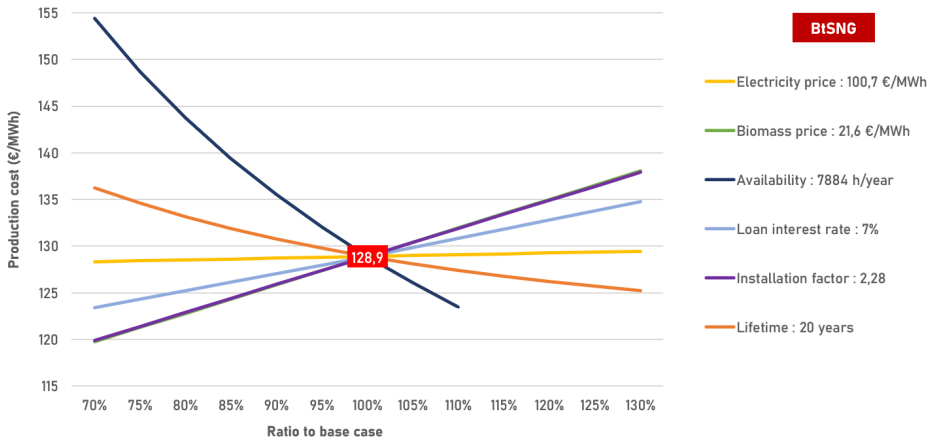


(b)

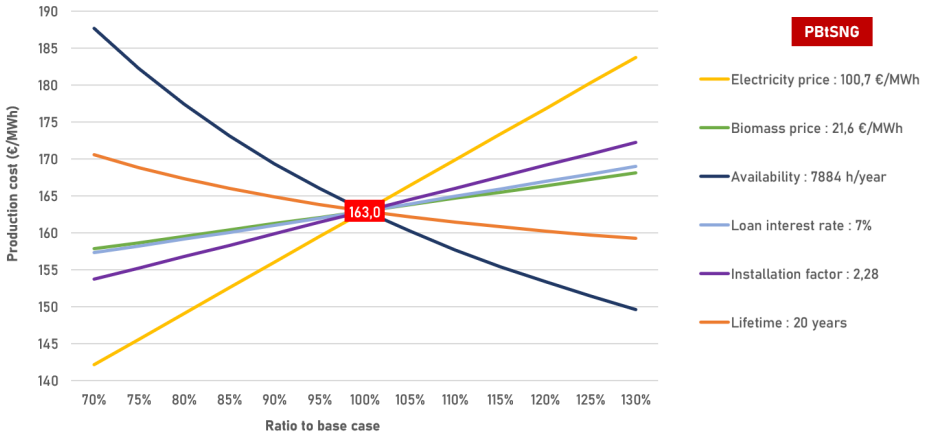


(c)

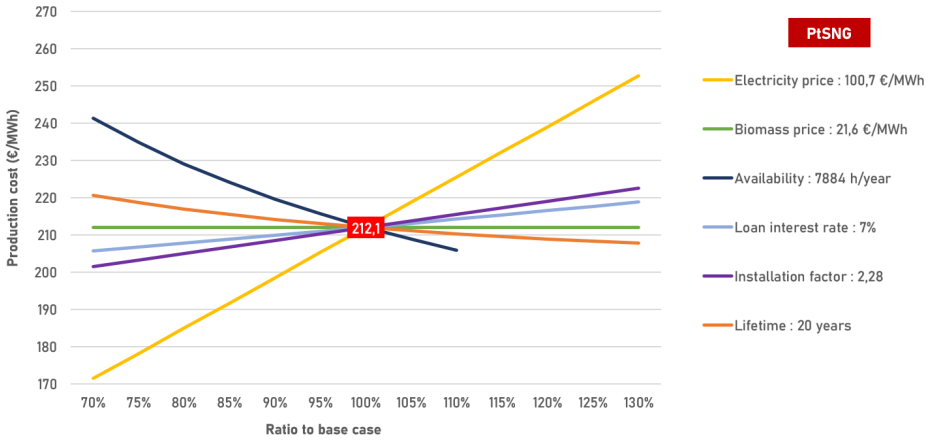
Figure 5: Sensitivity analysis on net production cost for BtMeOH (a), PBtMeOH (b), and PtMeOH (c) processes



(d)



(e)



(f)

Figure 6: Sensitivity analysis on net production cost for BtSNG (d), PBtSNG (e) and PtSNG (f) processes

13 GHG sensitivity analysis

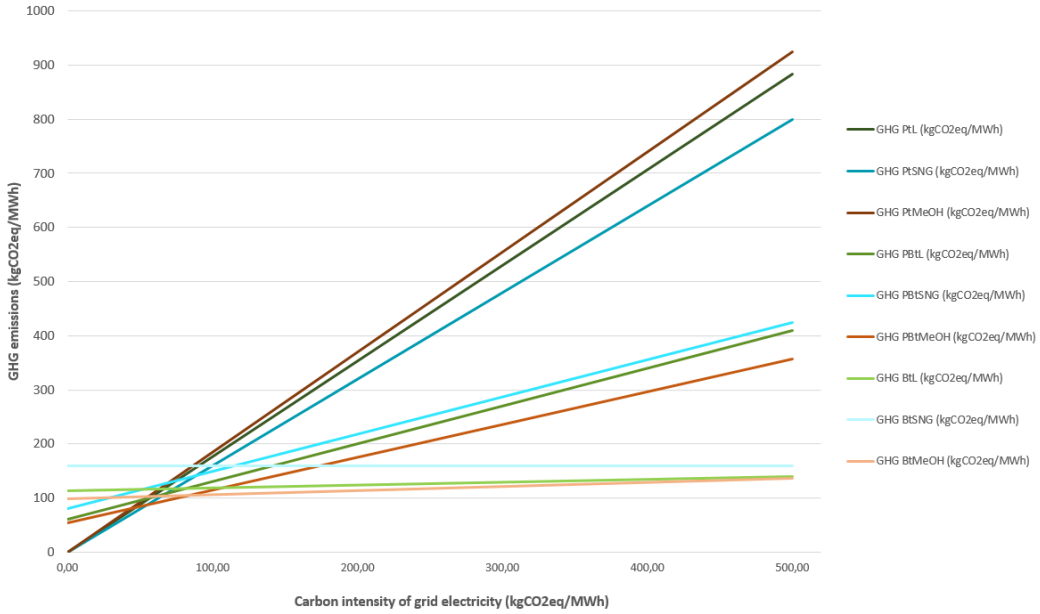


Figure 7: BtX, PBtX an PtX GHG emission sensitivity with respect to electricity carbon intensity

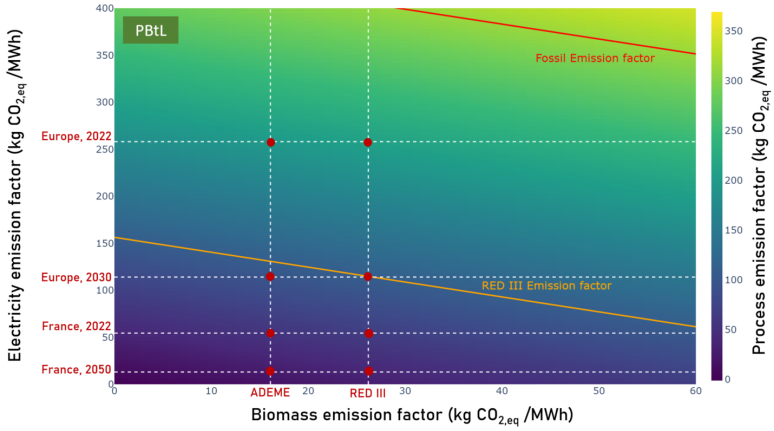


Figure 8: Cross-sensitivity analysis of PBtL GHG emissions depending on electricity and biomass emission factor. The red line represents the emission level of a fossil equivalent. The orange line represents the maximum permissible level of emissions to meet RED III requirements.

Table 8: GHG emissions of each process ($kgCO_2eq/MWh$) under different electricity mixes and biomass emission factor sources (ADEME and RED III). Values are coloured according to compliance with RED II sustainability thresholds.

Process	**GHG emissions of process ($kgCO_2eq/MWh$)**							
	Biomass EF source: ADEME				**Biomass EF source: RED III**			
	EU mix	**Nuclear**	**Renew. mix**	**CHP**	**EU mix**	**Nuclear**	**Renew. mix**	**CHP**
BtL	50.90	33.52	34.20	50.70	72.11	54.73	55.41	71.91
PBtL	245.54	21.67	30.41	242.95	256.92	33.05	41.79	254.33
PtL	575.04	9.71	31.78	568.50	575.04	9.71	31.78	568.50
BtSNG	95.29	95.29	95.29	95.29	112.29	112.29	112.29	112.29
PBtSNG	272.76	52.28	60.89	270.20	281.41	60.93	69.54	278.86
PtSNG	520.62	8.79	28.77	514.70	520.62	8.79	28.77	514.70
BtMeOH	53.65	29.29	30.24	53.37	72.08	47.72	47.72	71.80
PBtMeOH	213.60	19.29	26.87	211.35	223.78	29.47	29.47	221.53
PtMeOH	602.03	10.17	33.27	595.18	602.03	10.17	10.17	595.18

Note: Values represent process GHG emissions expressed per unit of final energy output ($kgCO_2eq/MWh$), calculated for different electricity mixes and biomass emission factor (EF) sources. Green values are **below the RED II sustainability threshold**, while red values **exceed it**. Thresholds correspond to at least **65% GHG reduction** for biofuels and e-biofuels, and **70% reduction** for e-fuels compared to fossil equivalents.
Thresholds applied: 109.24 (BtL, PBtL), 93.64 (PtL), 86.40 (BtSNG, PBtSNG, PtSNG), 86.94 (BtMeOH, PBtMeOH), and 74.52 $kgCO_2eq/MWh$ (PtMeOH).

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