



# Carbon-negative hydrogen: Exploring the techno-economic potential of biomass co-gasification with CO<sub>2</sub> capture

Carlos Arnaiz del Pozo<sup>a</sup>, Schalk Cloete<sup>b,\*</sup>, Ángel Jiménez Álvaro<sup>a</sup>

<sup>a</sup> Universidad Politécnica de Madrid, Spain

<sup>b</sup> SINTEF Industry, Norway

## ARTICLE INFO

### Keywords:

H<sub>2</sub> production  
Carbon capture and storage  
Biomass  
Membrane reactors  
Gasification

## ABSTRACT

The hydrogen economy is receiving increasing attention as a complement to electrification in the global energy transition. Clean hydrogen production is often viewed as a competition between natural gas reforming with CO<sub>2</sub> capture and electrolysis using renewable electricity. However, solid fuel gasification with CO<sub>2</sub> capture presents another viable alternative, especially when considering the potential of biomass to achieve negative CO<sub>2</sub> emissions. This study investigates the techno-economic potential of hydrogen production from large-scale coal/biomass co-gasification plants with CO<sub>2</sub> capture. With a CO<sub>2</sub> price of 50 €/ton, the benchmark plant using commercially available technologies achieved an attractive hydrogen production cost of 1.78 €/kg, with higher CO<sub>2</sub> prices leading to considerable cost reductions. Advanced configurations employing hot gas clean-up, membrane-assisted water-gas shift, and more efficient gasification with slurry vaporization and a chemical quench reduced the hydrogen production cost to 1.50–1.62 €/kg with up to 100% CO<sub>2</sub> capture. Without contingencies added to the pre-commercial technologies, the lowest cost reduces to 1.43 €/kg. It was also possible to recover waste heat in the form of hot water at 120 °C for district heating, potentially unlocking further cost reductions to 1.24 €/kg. In conclusion, gasification of locally available solid fuels should be seriously considered next to natural gas and electrolysis for supplying the emerging hydrogen economy.

## 1. Introduction

Increasing attention on the environmental footprint of fuels has led to accelerated research and development efforts in the direction of integrating practical and cost-effective carbon-free energy vectors in current energy systems. H<sub>2</sub> has great potential to decarbonize several industrial sectors, transportation, heating and power networks etc. as reflected in the IEA report [1]. Significant capital investments have taken place to generate “green H<sub>2</sub>” from excess renewable energy sources. However, drastic cost reduction in electrolyzer technology must still be achieved in order for this technology to be competitive with respect to current fossil fuel-based generation options. This is further accentuated when cost of handling intermittent H<sub>2</sub> production from solar and wind power is accounted for in the comparison, as highlighted by Cloete et al. [2].

Currently, fossil fuel-based processes are responsible of more than 95% of the total H<sub>2</sub> produced, with Steam Methane Reforming (SMR) from natural gas feedstocks providing 50% of the supply, as detailed by Voldstund et al. [3]. When hydrogen produced from fossil fuels integrate

carbon capture and storage (CCS) technologies, it is referred to as “blue H<sub>2</sub>”. The high energy penalty imposed by post and pre-combustion capture technologies result in a decreased competitiveness with high CO<sub>2</sub> avoidance costs of around 100 €/ton for pre-combustion capture with MDEA absorption, as demonstrated by Spallina et al. [4]. Khojasteh et al. [5] report that CCS integration increases the cost of H<sub>2</sub> by 40 to 100% relative to the unabated thermochemical conversion process. To address these economic concerns, novel technologies utilizing natural gas for H<sub>2</sub> production with integrated CO<sub>2</sub> capture have been the focus of research recently. Spallina et al. [4] present a detailed assessment of Membrane Assisted Chemical Looping Reforming (MA-CLR) and Fluidized Bed Membrane Reactor (FBMR) based H<sub>2</sub> plants yielding a negligible or even negative cost of CO<sub>2</sub> avoidance, at equivalent H<sub>2</sub> efficiencies of an SMR plant without CO<sub>2</sub> capture. Subsequently, Spallina et al. [6] carried out a proof of concept of a CLR dynamically operated packed bed reactor with a promising performance. On the other hand, Cloete et al. [7] present a Membrane Assisted Autothermal Reforming (MA-ATR) concept with a Levelized Cost of Hydrogen (LCOH) of 1.54 €/kg when the membrane is assumed to operate up to 800 °C. The cost of hydrogen production with CO<sub>2</sub> capture reported by

\* Corresponding author.

E-mail address: [schalk.cloete@sintef.no](mailto:schalk.cloete@sintef.no) (S. Cloete).

<https://doi.org/10.1016/j.enconman.2021.114712>

Received 12 July 2021; Accepted 27 August 2021

Available online 9 September 2021

0196-8904/© 2021 The Author(s). Published by Elsevier Ltd. This is an open access article under the CC BY license (<http://creativecommons.org/licenses/by/4.0/>).

**Nomenclature***Acronyms*

ASME	American Society of Mechanical Engineers
ASU	Air Separation Unit
BEC	Bare Erected Cost
BECCS	Bioenergy Carbon Capture and Storage
CA	CO <sub>2</sub> Avoided
CC	CO <sub>2</sub> Captured
CCS	Carbon Capture and Storage
CHH <sub>2</sub>	Combined Heat and Hydrogen
CLC	Chemical Looping Combustion
CLR	Chemical Looping Reforming
COT	Combustor Outlet Temperature
CPU	Cryogenic Purification Unit
CSTR	Continuous Stirred Tank Reactor
DH	District Heating
EOS	Equation of State
EPC	Engineering Procurement & Construction
FBMR	Fluidized Bed Membrane Reactor
FTR	Fired Tubular Reformer
GHG	Greenhouse Gases
GSR	Gas Switching Reforming
GT	Gas Turbine
HTS	High Temperature Shift
IEA	International Energy Agency
IEAGHG	International Energy Agency Greenhouse Gas R&D Program
LHV	Lower Heating Value
LP	Low Pressure

LT	Low Temperature
LTS	Low Temperature Shift
MAWGS	Membrane-Assisted Water-Gas Shift
MA-CLR	Membrane Assisted Chemical Looping Reforming
MA-ATR	Membrane Assisted Autothermal Reforming
NET	Negative Emissions Technology
NG	Natural Gas
NGCC	Natural Gas Combined Cycle
OC	Owner's Cost
PT	Project Contingency
PS	Process Contingency
PSA	Pressure Swing Adsorption
TC	Total Contingency
TIT	Turbine Inlet Temperature
TOC	Total Overnight Cost
TOT	Turbine Outlet Temperature
SE	Sorption Enhanced
SEA	Standardized Economic Assessment
SMR	Steam Methane Reforming
S/C	Steam to Carbon
RSC	Radiant Syngas Cooler
WGS	Water Gas Shift

*List of Symbols*

$\eta_{H_2}$	Hydrogen Efficiency
$\eta_{El}$	Electrical Efficiency
$C_{CO_2}$	CO <sub>2</sub> capture
$E_{CO_2}$	Specific Emissions
$\dot{m}$	Mass flow

Khan et al. [8] for a combination of chemical looping water splitting and combustion resulted in 1.68 \$/kg, significantly below the SMR benchmark with CO<sub>2</sub> capture. Nazir et al. [9] presents a techno-economic assessment of Gas Switching Reforming (GSR), which overcomes the technological challenges of interconnected fluidized beds at high pressure and operation with membranes at high temperature, reaching CO<sub>2</sub> avoidance costs ranging from 15 to 26.4 \$/ton of CO<sub>2</sub>. Alternative process intensification pathways such as sorption enhanced steam methane reforming (SE-SMR) evaluated by Yan et al. [10] present LCOH values between 1.90 and 2.53 £/kg. Lee et al. [11] provide a cost estimation of a lab scale sorption enhanced membrane reactor (SEMR) H<sub>2</sub> production plant of 3.04 \$/kg H<sub>2</sub>, indicating that such configuration achieves lower CO<sub>2</sub> emission rates relative to membrane and packed bed reactor designs. The different cost estimation bases and assumptions from these studies make it difficult to provide a consistent comparison between these advanced technologies. For perspective, the IEA report [12] indicates a H<sub>2</sub> cost from natural gas with conventional CO<sub>2</sub> capture of 1.73\$/kg (Europe, 2018), presenting a 34% levelized cost increase when CCS technology is integrated.

In parallel to natural gas feedstock plants, gasification technologies for H<sub>2</sub> and power co-production from solid fuels have also been gaining attention, as reflected in the IEAGHG report [13]. Such plants are an evolution of traditional pre-combustion CO<sub>2</sub> capture Integrated Gasification Combined Cycles (IGCC), where the generated syngas stream undergoes a series of water gas shift reactions and CO<sub>2</sub>/H<sub>2</sub>S removal by means of physical or chemical adsorbents for subsequent firing of a H<sub>2</sub> rich fuel in a Gas Turbine (GT) combustor. By integration of a Pressure Swing Adsorption (PSA) unit downstream the Acid Gas Removal plant, a substantial amount of the original heating value of the solid fuel can be retrieved as highly pure H<sub>2</sub> at pressurized conditions. The remaining PSA off-gas containing unconverted CO, CO<sub>2</sub> slip from the absorption unit and non-recovered H<sub>2</sub> is used for power generation in a Rankine

cycle upon combustion with air to raise HP steam, generating enough electricity to satisfy the plant auxiliary demand. A block flow diagram of this reference plant is provided in Fig. 1:

Alternatively to this well-known pathway for H<sub>2</sub> generation from solid fuels, Pd-based membranes investigated experimentally by Basile et al. [14] and originally proposed by Uemiya et al. [15] show great potential to enhance H<sub>2</sub> production efficiencies with low investment costs by reducing the scope of the syngas treating section of the plant, as studied by Jordal et al. [16]. The WGS reaction takes place along a membrane tube with H<sub>2</sub> diffusing across its surface, maximizing CO conversion and yielding a CO<sub>2</sub>/H<sub>2</sub>O stream in the retentate which, after water knock out, can be boosted to CO<sub>2</sub> delivery pressure with minimal power requirements. Through cryogenic purification of this stream [17], unconverted CO and unrecovered H<sub>2</sub> are retrieved and employed as fuel in a power cycle for electricity co-generation. Alongside membrane technology, high temperature contaminant removal from syngas is an opportunity to avoid the energy penalty derived from cooling and heating of syngas prior to the shift, as pointed out by Giuffrida et al. [18], while decoupling CO<sub>2</sub> and H<sub>2</sub>S removal, resulting both in capital investment and auxiliary power reductions relative to low temperature syngas treating options [19].

A particularly attractive aspect of gasification is the possibility to use biomass feedstock for achieving negative CO<sub>2</sub> emissions, critical for reaching net-zero emission targets in a cost-effective manner. However, biomass gasification is challenging due to aggressive molten slag (Higman [20]), tar formation at low temperatures (Sikarwar et al. [21]) and potential biomass supply issues in the volumes required for a large-scale hydrogen economy. Depending of the bioenergy source, the assumption of negative emissions with CO<sub>2</sub> capture technologies (BECCS) can be questionable, as pointed out by Fajardy et al. [22], due to emissions associated to cultivation, harvesting, transportation and processing. Co-gasification of a limited fraction of biomass with coal mitigates these

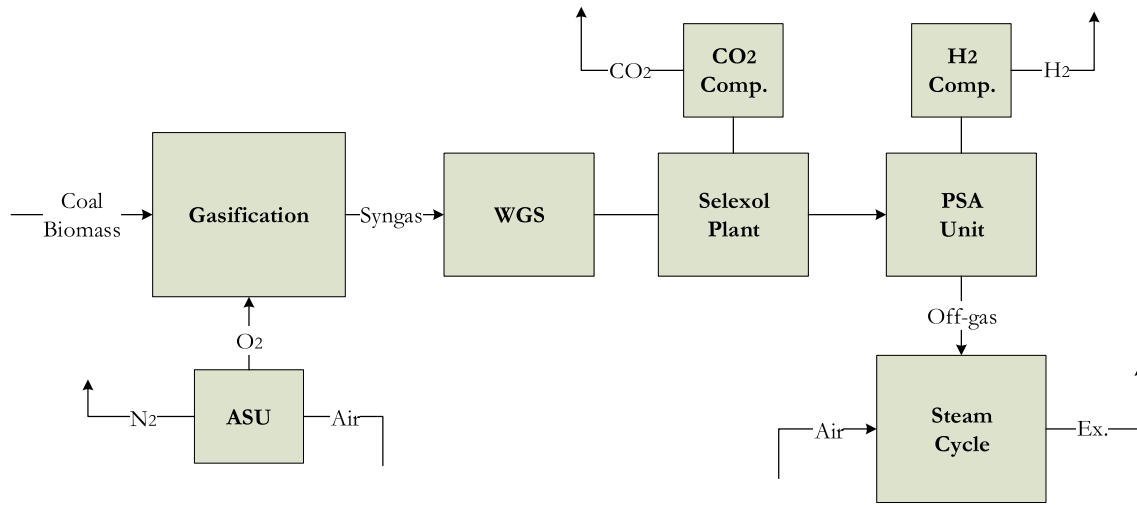


Fig. 1. Simplified block flow diagram of reference H<sub>2</sub> production plant. A detailed process flowsheet is provided in the Supplementary Material file.

challenges, while preserving the potential for negative emissions. Hetland et al. [23] highlight the advantages of co-firing biomass with coal in electricity generation plants with CCS. This is the route investigated in the current study, which presents a techno-economic assessment of a negative emissions technology (NET) for H<sub>2</sub> production employing an enhanced gasification system with a coal-biomass blend as feedstock, incorporating advanced technologies such as Pd-based membrane assisted water gas shift (MAWGS) reactors and hot gas clean up (HGCU) for syngas treating, conceptually represented in Fig. 2. This novel plant is benchmarked against the reference H<sub>2</sub> plant utilizing a GE gasifier with Selexol absorption for syngas treating and a PSA unit for H<sub>2</sub> purification, similar to the one presented in the IEAGHG report [13].

In the next section, a technological and modelling outline of the novel elements is presented, followed by detailed descriptions of the reference and advanced H<sub>2</sub> plant models. The energy and environmental metrics are defined, and the economic analysis methodology is presented. Subsequently, the model results for the different simulation cases are detailed, and the hydrogen production costs for the different configurations are reported. Sensitivity analyses to key process and economic assumptions are performed, to provide a holistic overview. Finally, the main conclusions from the study are presented, highlighting the principal impacts of the study in terms of the concept's

competitiveness and development challenges in the context of blue hydrogen production.

## 2. Methodology

In this section an outline of the process configurations for the reference and novel concepts is presented. Subsequently, the performance metrics in terms of energy, CO<sub>2</sub> emissions and economic analysis are outlined. The plants are modelled with the software Unisim Design R451 using Peng Robinson EOS for property calculation and ASME tables for steam. The Selexol unit in the reference H<sub>2</sub> plant is modelled using the thermodynamic parameters taken from Kapetaki et al. [24]. The MAWGS model is coded in Scilab using an in-house thermodynamic database, Patitug, for property estimation. The membrane reactor model is coupled to the stationary process flowsheet by means of a CAPE-OPEN unit operation. The energy balance relative error between the Scilab and Unisim platforms was below 0.1%.

### 2.1 Reference H<sub>2</sub> Plant

The reference plant model (Ref.) is developed based on the guidelines provided in an IEAGHG report [13], using a GE gasifier with an operating temperature of 1350 °C, and the maximum pressure indicated by Higman [20] (80 bar). A 95%mol O<sub>2</sub> stream is provided by a

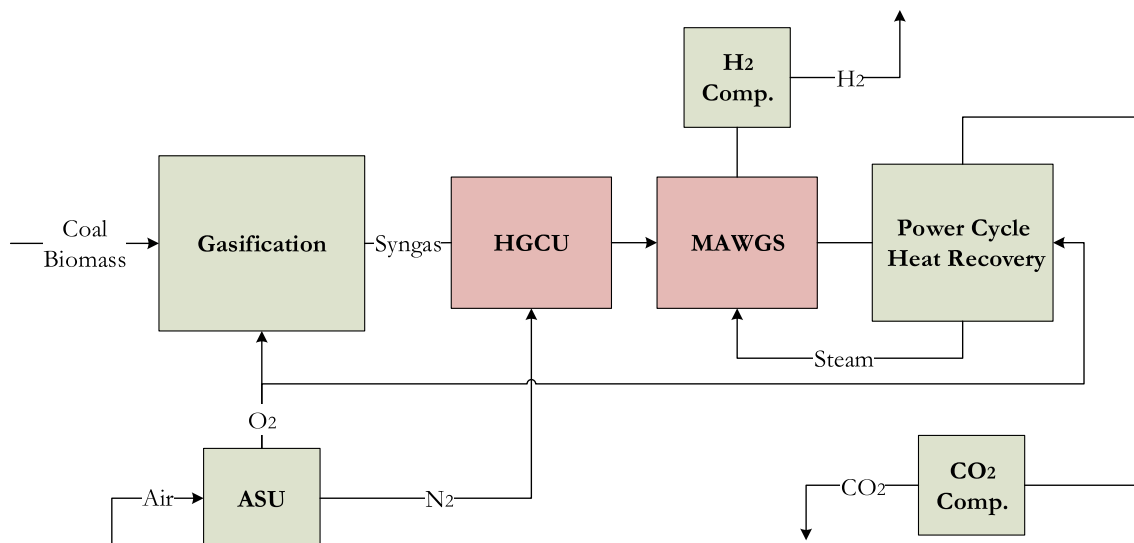


Fig. 2. Simplified block flow diagram of the advanced H<sub>2</sub> plant. A detailed process flowsheet is provided in the Supplementary Material file.

standalone low-pressure ASU consisting of a pumped liquid oxygen cycle, to avoid costly O<sub>2</sub> compressors and associated safety risks [25], to oxidize a blend of 30%w. of wood biomass [26] and Douglas Premium coal [27]. The feed slurry is assumed to be pre-heated to 250 °C with a resulting CGE of approximately 73%. A radiant syngas cooler (RSC) [28] is employed to generate HP superheated steam for syngas cooling to around 900 °C. A water quench achieves further syngas temperature reduction before it enters a saturator (modelled with three equilibrium stages) for fly ash and contaminant removal (NH<sub>3</sub>, chlorides etc.). The resulting steam to carbon ratio is approximately 3 prior to a sour shift, composed of two intercooled adiabatic reactors operating at around 200–300 °C, achieving a final CO conversion of approximately 98%. The shifted syngas is routed to a Selexol plant after heat recovery to selectively remove H<sub>2</sub>S and CO<sub>2</sub>. Given the high partial pressure of the latter species, a removal efficiency of around 95% is reached. Emissions originate from unconverted CO, CO<sub>2</sub> slip from the absorption column and a small fraction of CH<sub>4</sub> present in the syngas, yielding an overall capture rate of 93%. The H<sub>2</sub> rich syngas is routed to a Pressure Swing Adsorption Unit (PSA) modelled with a recovery correlation presented by Nazir et al. [29], assuming a H<sub>2</sub> purity of 100%. The PSA off gas pressure for adsorbent regeneration is selected in such a way that the heat released upon combustion with an air stream in a dedicated boiler raises sufficient steam (110 bar and 550 °C) to achieve an electrically sufficient scheme upon expansion in a steam turbine, assuming a condenser pressure of 0.048 bar. This resulted in a PSA off-gas pressure of 1.85 bar and 90.5% H<sub>2</sub> recovery. Other steam sources such the RSC and WGS heat recovery units are integrated with the steam cycle to maximize power production. Further modelling assumptions of the plant units as well as a detailed process diagram and corresponding stream summary can be found in the [Supplementary Material](#).

### 2.1. Advanced H<sub>2</sub> plants

Three advanced plant configurations are investigated: two models are designed assuming that all power consumption must be satisfied with electricity produced internally, and evaluate the trade-offs between different power cycles implemented to accomplish this. The final case assumes that some electricity can be imported outside battery limits, to maximize H<sub>2</sub> yield. A description of the advanced technology elements incorporated in each of these concepts as well as detailed process diagrams and stream summaries, together with further modelling assumptions are provided as [Supplementary material](#).

1. Wnet0 - ST: An electrically self-sufficient plant that employs a GE gasifier and HGCU treating to deliver syngas to a MAWGS reactor. Power is produced through oxy-combustion of the MAWGS retentate stream with pure O<sub>2</sub> from the ASU to generate additional steam for a steam turbine. Thus, this configuration produces electricity in an equivalent way to the Ref. model, with a significant portion of the steam delivered by an RSC.
2. Wnet0 - GT: In this process concept, the benefits of a higher CGE from the E-gas gasifier (studied by Gräbner et al. [30]) with a second stage chemical quench instead of an RSC and slurry vaporization (experimentally proven by Aiuchi et al. [31]) are reaped by implementing a more efficient power cycle, i.e. a gas turbine (GT), for electricity generation. Fuel for the GT is obtained through cryogenic purification of the retentate stream, as shown by Jordal et al. [16].
3. MaxH<sub>2</sub>: This configuration also integrates a two stage E-gas gasifier with slurry vaporization to maximize CGE. The power cycle is omitted and some electricity imports are necessary to satisfy the plant auxiliary power demand, designing the MAWGS to maximize H<sub>2</sub> extraction. The low heating value syngas in the retentate stream is combusted in the same way as in the Wnet0 – ST model, generating steam for the sweep after expansion in a back-pressure turbine, which somewhat reduces the net power required.

Some considerations should be made regarding the design of these advanced H<sub>2</sub> plants. Through the water quench cooling of the syngas coming out of the gasifier (from the RSC of the GE gasifier and the chemical quench for the E-gas depending on the case), the steam to carbon ratio is adjusted to a suitable value of 1.9, to carry out the WGS reaction downstream without catalyst deterioration. Prior to this step the syngas is fed to a hot gas desulphurization unit (HGCU) operating in the range between 400 and 500 °C to remove sulphur species and other contaminants. Clean syngas is subsequently cooled in a HP steam evaporator to approximately 320 °C and converted in a high temperature shift (HTS) stage, reaching around 75% CO conversion (assumed at equilibrium) and resulting in an adiabatic temperature rise to approximately 500 °C. The downstream effluent is further cooled producing HP superheated steam and sent to the MAWGS reactor, achieving overall CO conversion values above 90% and high H<sub>2</sub> recoveries per unit of area due to the large driving force, as the syngas stream is fed at around 70 bar. The MAWGS is operated with a counter-current steam sweep, which maximises both CO conversion (by quenching the exothermic reaction) and H<sub>2</sub> recovery (by decreasing its partial pressure), while obtaining a pressurized product stream, minimizing H<sub>2</sub> compression costs and auxiliary consumption. The inlet temperature of the syngas stream is controlled in each case by the amount of steam generation to limit the maximum temperature in the retentate side of the MAWGS to 500 °C.

Through a techno-economic optimization, it was determined that the permeate pressure which resulted in the lowest LCOH was approximately 15 bar, leading to a H<sub>2</sub> compression train comprising two intercooled stages for a delivery pressure of 60 bar, as suggested in the European Hydrogen Backbone site [32]. By increasing the *n*° of membrane tubes, larger sweep pressures could be imposed, but the reduction in H<sub>2</sub> compression costs did not outweigh the capital cost increase of the MAWGS reactor. On the other hand, decreasing the permeate pressure beyond the optimal value led to an additional compression stage to avoid a large temperature rise, resulting in higher overall costs. The amount of steam that was employed as sweep in the membrane for the Wnet0 - ST and Wnet0 – GT plant concepts was manipulated to reach a net electrical consumption of zero. Small variations from the optimal pressure for the nearest 500 membrane tubes were made to the models to reach the specified temperature approaches in the heat recovery network, which had small impacts in H<sub>2</sub> compression costs comparatively. Heat rejection for all cases was simulated with a cooling water tower (approximated to 3 equilibriums stages) to reach a cooling water temperature of 20 °C, and a process stream temperature after heat rejection of 25 °C.

For the Wnet0 – ST and MaxH<sub>2</sub> cases, the retentate outlet, with a small fraction of the heating value from the original syngas is combusted with pure O<sub>2</sub> produced by the ASU. Combustible species consist of unconverted CO, unrecovered H<sub>2</sub> and the CH<sub>4</sub> produced in the gasifier. Since their concentration can be low due to the large heating value extraction in the form of H<sub>2</sub> upstream, in this work it is considered that oxidation can be enhanced by means of a metallic oxygen carrier (e.g., Cu/CuO or Ni/NiO) in a fluidized bed, to ensure complete conversion of the reactants. After heat recovery and cooling to ambient temperatures, traces of water which have not condensed in a knock-out vessel are removed in a dehydration unit resulting in a CO<sub>2</sub> stream with a purity close to 97%mol (the remaining consists of N<sub>2</sub> and Ar from the ASU oxidant streams). Condensed water with some dissolved CO<sub>2</sub> is pumped, reheated and recycled to the syngas quench to achieve 100% CO<sub>2</sub> capture. The outlet gas stream is boosted above supercritical pressure, and further pumped to 150 bar for transport and storage.

On the other hand, for the Wnet0 – GT case requiring a suitable fuel for a GT, the retentate stream oxy-combustion is not carried out: after heat recovery, water knock out and dehydration, the stream is purified in a double-flash cryogenic purification unit (CPU), similar to the one presented by Campanari et al. [17] and implemented by Jordal et al. [16]. The purge stream contains most of the combustible species and is sent to the GT combustor, leading to some unavoidable emissions

originating from the CO, CH<sub>4</sub> and CO<sub>2</sub> present in this stream. The resulting CO<sub>2</sub> purity values were consistent with the guidelines recommended for transport and storage [33], although small fractions of CO, CH<sub>4</sub> and H<sub>2</sub> were present due to the lack of complete separation in the CPU, as opposed to the previous cases where the only source of impurities are inert N<sub>2</sub> and Ar from the oxidant streams delivered by the ASU.

## 2.2. Plant performance indicators

In this section, the metrics in terms of energy, environmental and economic performance are defined. The H<sub>2</sub> production plant concept can be schematically represented as shown in Fig. 3:

### 2.2.1. Energy and environmental

The energy performance of the plant is defined attending to the total fuel heating value input, the heating value of the H<sub>2</sub> product and the net electricity demand/production of the plant ( $\dot{W}_{net}$ ). In this work, the same efficiency weight is given to electrical and hydrogen efficiencies, based on the assumption that hydrogen prices will be similar to electricity prices in a future hydrogen economy. The hydrogen ( $\eta_{H_2}$ ), electrical ( $\eta_{El}$ ), and overall net ( $\eta_{net}$ ) efficiencies are defined as shown in Eqs. (1)–(3), respectively:

$$\eta_{H_2} = \frac{\dot{m}_{H_2} LHV_{H_2}}{\dot{m}_{coal} LHV_{coal} + \dot{m}_{bio} LHV_{bio}} \quad (1)$$

$$\eta_{El} = \frac{\dot{W}_{net}}{\dot{m}_{coal} LHV_{coal} + \dot{m}_{bio} LHV_{bio}} \quad (2)$$

$$\eta_{net} = \eta_{H_2} + \eta_{El} \quad (3)$$

CO<sub>2</sub> capture ( $C_{CO_2}$ ) is defined as the ratio between mass flow from the CO<sub>2</sub> pump discharge and the total CO<sub>2</sub> generated in the plant, both from biomass and coal origin (Eq. (4)). On the other hand, the net CO<sub>2</sub> capture rate ( $C'_{CO_2}$ ) discounts from the denominator the CO<sub>2</sub> from biomass origin, returning values above 100% and negative specific emissions (Eq. (5)). The fraction of CO<sub>2</sub> originated from biomass is directly determined with the ultimate analysis on a moisture and ash free basis of the coal and biomass and their weight proportion in the blend. Finally, the specific emissions ( $E_{CO_2}$ ) are determined with the total heating value of the H<sub>2</sub> product (Eq. (6)) and the net emissions flow rate. CO<sub>2</sub> emissions involved in electricity imports are neglected on the assumption that the electricity sector will be largely decarbonized by the time that these negative-emission plants are deployed at scale.

$$C_{CO_2} = \frac{\dot{m}_{CO_2,capt.}^{bio} + \dot{m}_{CO_2,capt.}^{coal}}{\dot{m}_{CO_2,capt.}^{coal} + \dot{m}_{CO_2,emit.}^{coal} + \dot{m}_{CO_2,capt.}^{bio} + \dot{m}_{CO_2,emit.}^{bio}} \quad (4)$$

$$C'_{CO_2} = \frac{\dot{m}_{CO_2,capt.}^{bio} + \dot{m}_{CO_2,capt.}^{coal}}{\dot{m}_{CO_2,capt.}^{coal} + \dot{m}_{CO_2,emit.}^{coal}} \quad (5)$$

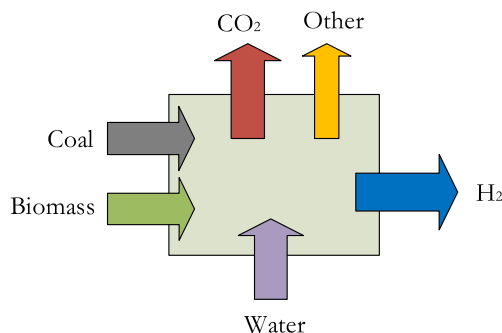


Fig. 3. Schematic representation of the H<sub>2</sub> plants.

$$E_{CO_2} = \frac{\dot{m}_{CO_2,emit.}^{coal} - \dot{m}_{CO_2,capt.}^{bio}}{\dot{m}_{H_2} LHV_{H_2}} \quad (6)$$

### 2.2.2. Economic analysis

The economic assessment methodology employed for the capital cost and operating & maintenance costs estimation and cash flow analysis is similar to the approach followed by the authors in a previous work [34]. The objective is to determine the levelized cost of hydrogen (LCOH), the price at which the product must be sold to arrive at a net present value (NPV) of 0 at the end of plant operation. NPV is determined through the discounted annual cash flow rates, which incorporate plant revenues, fixed and variable operating costs, capital costs and emissions taxes.

A Standardized Economic Assessment (SEA) Tool has been specifically developed by the authors in MS Excel with the aim to provide a transparent reproducible methodology for economic evaluation of chemical and energy plants [35]. A comprehensive user guide of this tool is available [36]. The full economic assessments of all the plants evaluated in this study are available for download [37], where details of the capital cost and operational expenditure estimations of each plant are provided. The most important function of the tool is to provide reliable and convenient estimates for the Bare Erected Costs (BEC) of the different plant units. For this purpose, two estimation approaches are followed. For well-known units, equipment lists from the process simulation are elaborated and each unit's cost is assessed by means of correlations from Turton et al. [38]. For less conventional units, a cost capacity estimate is employed similarly to Anantharaman et al. [27]. All unit estimates are adjusted from the source cost basis to a specified target cost basis by means of location factors to account for differences in material and labour costs between regions provided by an IEAGHG report [39], exchange rates for currency and Chemical Engineering Plant Cost Index (CEPCI) [40] for different cost-year estimates. The target basis for the present economic evaluation is defined in Table 1:

As the main source for scaling cost estimation, the costs reported in the NETL report [41] were used, since both gasification technologies investigated in this work are presented in this assessment, to provide a consistent comparison. Given the disparity of cost estimations for gasification-based plant units available in literature [13,27,41,42], a dedicated sensitivity to the gasifier cost is performed. The SEA tool also presents the possibility to benchmark user estimates with reference values adjusted to the target cost basis. As a supplementary cost estimation reference the IEAGHG report [13] presents a detailed description and economic parameters for the evaluation of a GE gasification-based plant for H<sub>2</sub> production.

To reach the Total Overnight Costs (TOC) of the evaluated plant, the methodology described in Table 2 was applied. Engineering Procurement and Construction costs were taken to be 10%, within the range suggested by Roussanaly et al. [43]. The project contingency (PC) and owner's costs (OC) were taken to add up to 35%, slightly higher than the value assumed in previous studies by the authors (Szima et al. [34]). This high value relative to the recommendations given by Anantharaman et al. [27] is considered to reflect uncertainties related to construction of plants involving gasification technologies. Process contingencies are accounted for to reflect a relatively lower degree of technological maturity of specific units with respect to commercial and well-known processes. For the advanced plants, both the HGCU and MAWGS units assume a PC of 30%, whereas for the unconventional steam-power cycle, using fluidized boilers to carry out oxy-combustion, a 10% PC was taken. Furthermore, the advanced features considered for

Table 1  
Target basis for economic assessment.

Location	Western Europe
Cost Year Basis	2018
Currency	€



**Table 2**  
Cost estimation methodology for the plant TOC.

Component	Definition
Bare Erected Cost (BEC)	Sum of all BEC costs of the units adjusted to the target basis in Table 1
Engineering Procurement and Construction (EPC)	10% of BEC
Process Contingency (PC)	0–30% of BEC
Project Contingency (PT)	20% of (BEC + PS)
Total Plant Costs (TPC)	BEC + EPC + PS + PT
Owners Costs (OC)	15% of TPC
Total Overnight Costs (TOC)	TPC + OC

the E-gas gasification (high operating pressure and slurry vaporization), justify a 10% PC applied to this unit. Finally, for the Wnet0 – GT plant, a PC of 10% was applied to the CO<sub>2</sub> CPU.

The cash flow analysis methodology assumes a baseload H<sub>2</sub> production plant with the economic evaluation baseline presented in Table 3. A dedicated sensitivity to the capacity factor and discount rate was carried out, the latter being a critical variable in the economic metric for competitiveness of capital-intensive low-carbon plants, as highlighted by Hirth et al. [44].

The main assumptions for fixed and variable operating and maintenance costs are summarized in Table 4. Sensitivities of the LCOH to the fuel costs, CO<sub>2</sub> tax, electricity price, and membrane costs are presented in the following sections. CO<sub>2</sub> tax levels are of particular interest in this study since, for a given biomass price, the income resulting from negative emissions can at a certain CO<sub>2</sub> price outweigh the higher costs of biomass relative to coal (on an energy basis).

### 3. Results

This section presents the H<sub>2</sub> plant model results, with a detailed energy breakdown for the cases considered, followed by the energy & CO<sub>2</sub> environmental metrics defined earlier. Then, the results of the economic analysis is shown, followed by a sensitivity analysis of the LCOH to the key economic assumptions.

#### 3.1. Energy & environmental results

The energy breakdown results of the configurations discussed in the previous section are presented in Table 5; the energy and environmental metrics defined in Section 2.3.1 are also detailed below. These models assume a co-gasification of biomass with coal with a fraction of 30% on a mass basis, with a total heat input of 1254.2 MW, analogously to the plants evaluated in the IEAGHG report [13].

The Ref. model shows the lowest H<sub>2</sub> efficiency (59.3%), due to the large auxiliary consumption incurred upon by the CO<sub>2</sub> compression and Selexol units. This result is around 4%-points above the values reported in the IEAGHG report [13], due to a lower gasifier performance relative to this study and because some electricity is produced in excess, reducing therefore the amount of H<sub>2</sub> product. A reduced capture rate results from CO<sub>2</sub> slip, CH<sub>4</sub> and unconverted CO in the absorption system, which are combusted in a boiler for power generation in a steam cycle. The Wnet0 – ST model integrates advanced elements such as HGCU and MAWGS, considerably decreasing the overall auxiliary consumption, as the CO<sub>2</sub> is obtained at high pressure, leading to a 3.6%-points H<sub>2</sub>

**Table 3**  
Cash flow analysis assumptions [13]

Component	Value
Economic Lifetime	25 years
Construction Period	4 years
Discount Rate	8%
Capacity Factor 1st year	65%
Capacity Factor 2nd year	85%

**Table 4**  
Fixed and variable operating & maintenance cost assumptions.

Fixed O&M	
Maintenance	2.5% of TOC [13]
Insurance & Taxes	1.0 % of TOC [13]
Labour	60,000 €/p-y [13]
Variable O&M	
Cooling water make-up	0.325 €/m <sup>3</sup> [27]
Process water costs	6 €/m <sup>3</sup> [27]
Selexol make-up	5000 €/ton [27]
WGS catalyst	16,100 €/m <sup>3</sup> [9]
Oxygen carrier	15\$/m <sup>3</sup> [45]
ZnO cost	25,230 \$/m <sup>3</sup> [46]
Ash disposal cost	9.73 €/m <sup>3</sup> [19]
Membrane replacement	6000 €/m <sup>2</sup> [47]
CO <sub>2</sub> transport and storage	10 €/ton
CO <sub>2</sub> emissions tax	50 €/ton
Electricity price	60 €/MWh
Fuel Costs	
Biomass	100 €/ton (6.1 €/GJ) [13]
Coal	2.5 €/GJ [13]

**Table 5**  
Energy and environmental results.

Model/Item	Ref.	Wnet0 – ST	Wnet0 – GT	MaxH <sub>2</sub>
H <sub>2</sub> LHV (MW)	744.2	788.2	846.8	915.6
H <sub>2</sub> flow (kg/s)	6.20	6.57	7.06	7.63
<i>Auxiliary Electricity Consumption</i>				
ASU (MW)	58.3	62.8	42.4	51.8
H <sub>2</sub> compressor (MW)	0.0	17.3	18.5	17.3
CO <sub>2</sub> compressor (MW)	25.3	2.8	7.9	3.0
Pumps (MW)	3.2	2.9	1.9	2.0
Gasifier aux. (MW)	3.4	3.4	3.4	3.4
Syngas treating aux. (MW)	25.3	0.7	0.7	0.7
Heat rejection (MW)	2.1	2.4	1.3	1.7
Other (MW)	0.7	0.0	0.0	0.0
<i>Electricity Generators</i>				
Gas turbine (MW)	0.0	0.0	54.0	0.0
Steam turbine (MW)	119.0	93.0	22.4	29.5
<i>Energy Efficiency</i>				
$\eta_{H_2}$ (%)	59.3	62.9	67.5	73.0
$\eta_{El}$ (%)	0.1	0.1	0.0	-4.0
$\eta_{net}$ (%)	59.4	62.9	67.5	69.0
<i>CO<sub>2</sub> Emissions</i>				
$C_{CO_2}$ (%)	93.8	100.0	91.7	100.0
$C'_{CO_2}$ (%)	121.0	129.0	118.3	129.0
$E_{CO_2}$ (kg/MWh-H <sub>2</sub> )	-95.7	-124.9	-72.3	-107.8

efficiency improvement. Increased ASU size and H<sub>2</sub> compression duty offset to a small extent these power reductions. However, for the Wnet0 – GT model utilizing a more efficient power cycle, the higher CGE of the gasifier leads to a further increase of H<sub>2</sub> efficiency of 4.6%-points. Overall, the auxiliary consumption is reduced because of a smaller ASU, despite the small increase in duty associated to the CO<sub>2</sub> CPU. Nonetheless, a significant reduction in capture ratio due to the purge of the CPU unit, which is employed as GT fuel, is observed. Finally, the MaxH<sub>2</sub> case presents the highest H<sub>2</sub> efficiency (13.7%-points above the Ref. model), albeit at a net power consumption of 4% (relative to the heat input) which must be provided through electricity imports. Given the large CO<sub>2</sub> fraction derived from the biomass fuel, all models present negative specific emissions and net capture rates above 100%.

#### 3.2. Economic assessment results

This section presents the economic results of the base case models,

followed by a sensitivity study of to the main process and economic assumptions. The effect of varying the co-gasified biomass mass fraction is investigated and finally the maximum economic potential of the plant considering district heating (DH) hot water sales is determined.

### 3.2.1. Base case results

A summary of the base case results in terms of specific capital investment can be seen in Fig. 4. Full details can be accessed in the SEA tool datasheets of each model for detailed values of the unit's cost estimation [37]. Overall, it can be seen that the most capital-intensive items are the ASU and gasification units, which in all cases represent more than 50% of the total capital cost. Notably, the Ref. plant with GE gasification, WGS unit, Selexol plant and PSA purification presents the highest specific capital investment, as a result of the large costs of these units together with the lowest H<sub>2</sub> production efficiency. When integrating advanced elements such as HGCU and MAWGS reactors, with an oxy-boiler for steam generation in Wnet0 -ST case, the cost increase due to H<sub>2</sub> compression is largely outweighed by the cost reductions provided by these items and the somewhat lower cost of the power cycle. The auxiliary power demand is to an important extent curtailed due to the avoidance of the Selexol plant and minimization of the CO<sub>2</sub> compression train, leading to a smaller power cycle. When a more efficient power generation system is implemented (Wnet0 - GT), efficiency benefits from higher CGE in the E-gas gasifier with slurry vaporization are translated into higher H<sub>2</sub> production efficiency and in parallel, cost reductions are achieved both in the ASU (lower O<sub>2</sub> consumption as no oxy-firing is implemented) and power cycle (GT). This results in an even lower overall specific capital investment for this case, despite the notable increase in cost of the CO<sub>2</sub> compression section, requiring a CO<sub>2</sub> CPU. Finally, for the MaxH<sub>2</sub> case where the power generation for auxiliary consumption is neglected assuming some electricity imports, the slight increase in the ASU, heat recovery section (oxy-boiler) and larger MAWGS reactor are outweighed by the large H<sub>2</sub> efficiency enhancement. Overall, the cost of the power cycle and heat recovery section (only generating some power with a backpressure steam turbine) is reduced relative to the previous case. Analogously to the Wnet0 - ST case, the cost associated to CO<sub>2</sub> compression is minimized, as a highly pure and pressurized CO<sub>2</sub> stream is obtained after water knock-out from the oxy-combusted retentate stream.

The specific operating & maintenance costs are presented in Fig. 5,

for each of the cases. Notably, the plants with lower CO<sub>2</sub> capture (Ref. & Wne0-GT) present lower revenues from negative emissions while the plant MaxH<sub>2</sub> shows a higher variable operating cost due electricity imports. Relative to the Ref. plant, the configurations employing HGCU and MAWGS units show a slightly higher VOM due to the costs associated to membrane and sorbent replacement, as well as somewhat higher process water costs due to a larger H<sub>2</sub> production. FOM contribution to the overall operating costs are smaller for the advanced plants given the lower estimated capital investments.

The LCOH distribution for all the base cases is shown in Fig. 6. As discussed, CO<sub>2</sub> tax creates a revenue for the plant (due to negative CO<sub>2</sub> emissions). Based on these results, the Wnet0 - ST plant can produce H<sub>2</sub> at a cost of 0,15 €/kg (8.6%) cheaper than the Ref. plant. This reduction is extended to 0,22 €/kg (12.2%) when a more efficient gasification and power generation systems are employed (Wnet0 - GT). As observed earlier, LCOH reduction due to CO<sub>2</sub> tax revenues for this plant and the Ref. model is comparatively smaller to the rest, due to lower CO<sub>2</sub> capture. When electricity imports are allowed (MaxH<sub>2</sub>), the combination of enhanced H<sub>2</sub> production and lower capital expenditure yield a LCOH which is 0,28 €/kg (15.7%) cheaper than the Ref. plant. For this last case, the notable increase in VOM costs results from electricity imports. The relatively low costs of H<sub>2</sub> achieved for the plants of this study, despite their high capital intensity, arise from the benefits of the economies of scale given the large H<sub>2</sub> production capacities, compared to the smaller natural gas based plants [12].

### 3.2.2. Sensitivity to economic parameters

Given the methodological and market uncertainties in this assessment, several sensitivity analyses of the LCOH to key assumptions of the evaluation were carried out. These cases, which are presented in Fig. 7, can be classified in three main blocks: operational expenditure (cost of biomass and coal, CO<sub>2</sub> tax, electricity price), capital costs (gasifier cost, membrane costs), and market factors (capacity factor, discount rate).

Even though biomass is more than twice as expensive per unit of energy as coal, the sensitivity to the biomass price is lower than to the coal price because it provides only 22% of the LHV energy input to the plant. However, this flexible fuel plant could still benefit from adjusting the coal/biomass ratio in response to price signals of these two fuels. The MaxH<sub>2</sub> plant shows a moderate sensitivity to the electricity price, but it remains cheaper than the electrically self-sufficient plants even at an

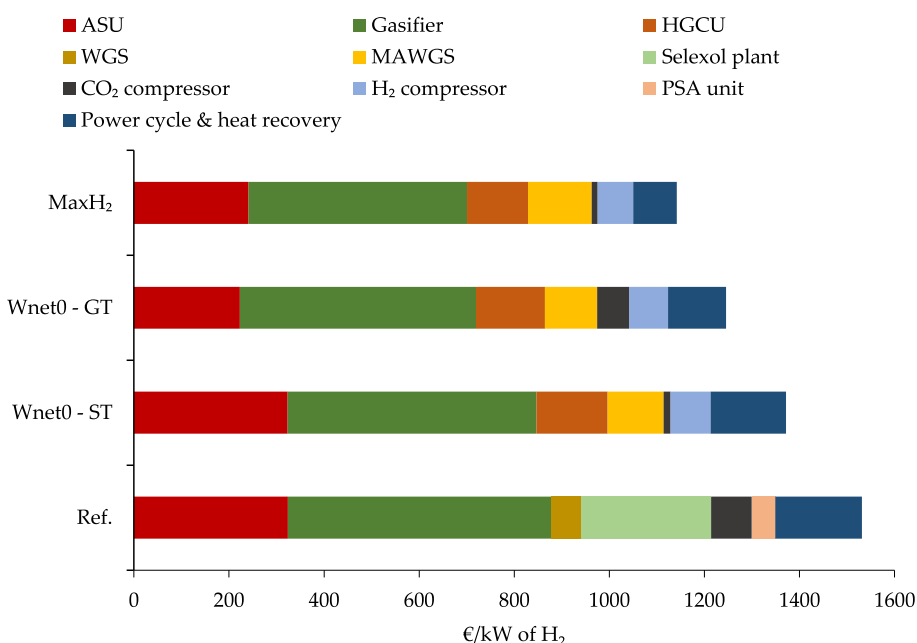


Fig. 4. Specific capital investment for the base case models.

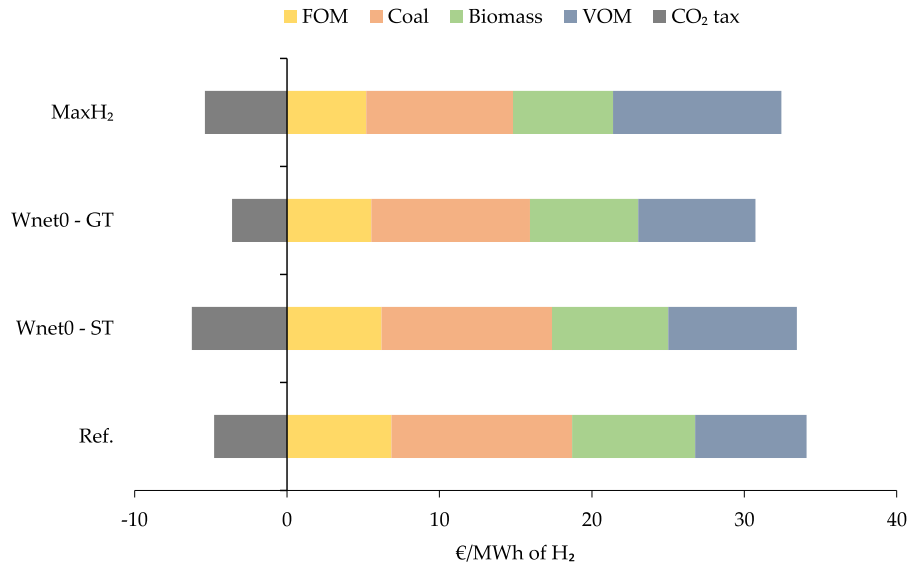


Fig. 5. O&M costs per MWh of H<sub>2</sub> for the base case models.

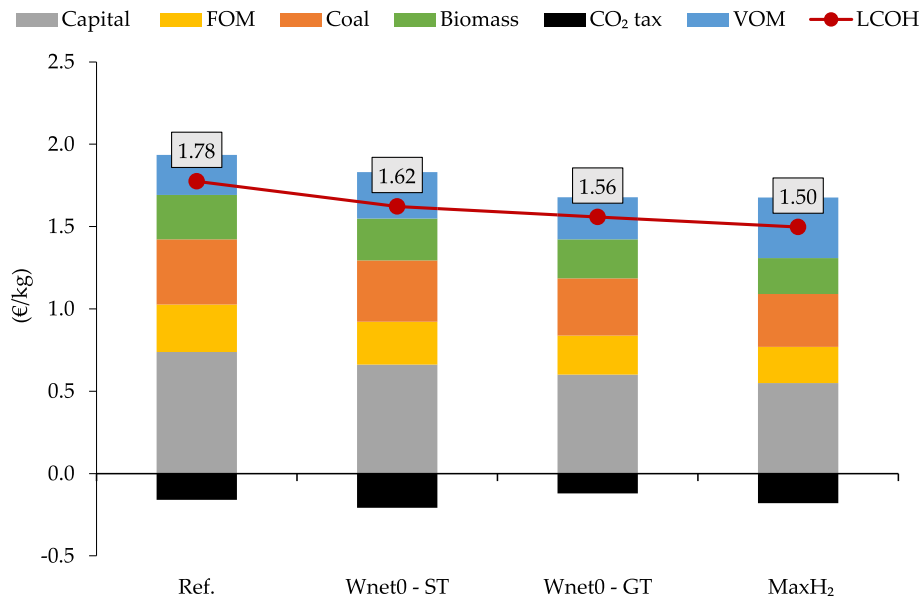


Fig. 6. LCOH results for the base case models.

electricity price of 80 €/MWh. Higher CO<sub>2</sub> prices have a large positive effect on all four plants due to their negative emissions, as will be discussed later. However, plants with capture rates below 100% (Ref. & Wnet0 – GT) present a comparatively lower sensitivity to CO<sub>2</sub> taxes, with the Wnet0 – GT plant resulting in a higher LCOH compared to the Wnet0 – ST plant at CO<sub>2</sub> prices of 100 €/ton, despite employing a more efficient power generation cycle. If electric self-sufficiency is important in an environment of high CO<sub>2</sub> taxes, a cost-effective alternative could be to combust some of the H<sub>2</sub>/steam mixture from the membranes in the MaxH<sub>2</sub> case using a power cycle configuration similar to the Wnet0 – GT plant.

The discount rate has a significant effect on these relatively capital-intensive plants. The MaxH<sub>2</sub> plant shows the lowest sensitivity due to its lower specific capital cost (Fig. 4). A significant sensitivity to the capacity factor is also observed. Unlike thermal power plants that are increasingly required to balance variable renewables, thermochemical hydrogen plants are likely to operate maximizing the operating hours

per year. The gasifier is most likely to dictate the achievable capacity factor in practice.

Being the costliest component in the plant, the relatively uncertain gasifier bare erected cost also has a significant effect on the LCOH. In general, a ± 30% cost variation in the gasifier BEC has roughly the same effect as a ± 30% cost variation in the coal price. Given the importance of gasifier technology both for plant capital costs and capacity factors, optimization of this component is key for the success of these negative-emission hydrogen plants. Finally, varying the membrane cost over an order of magnitude had a smaller effect than the ± 30% gasifier cost variation, although the MaxH<sub>2</sub> case with a greater membrane surface area to maximize H<sub>2</sub> permeation shows a slightly steeper sensitivity. Various uncertain membrane performance metrics can also influence this sensitivity, including lifetime, permeability (influencing the required membrane surface area), and cost recovery (re-using some parts of the spent membranes).



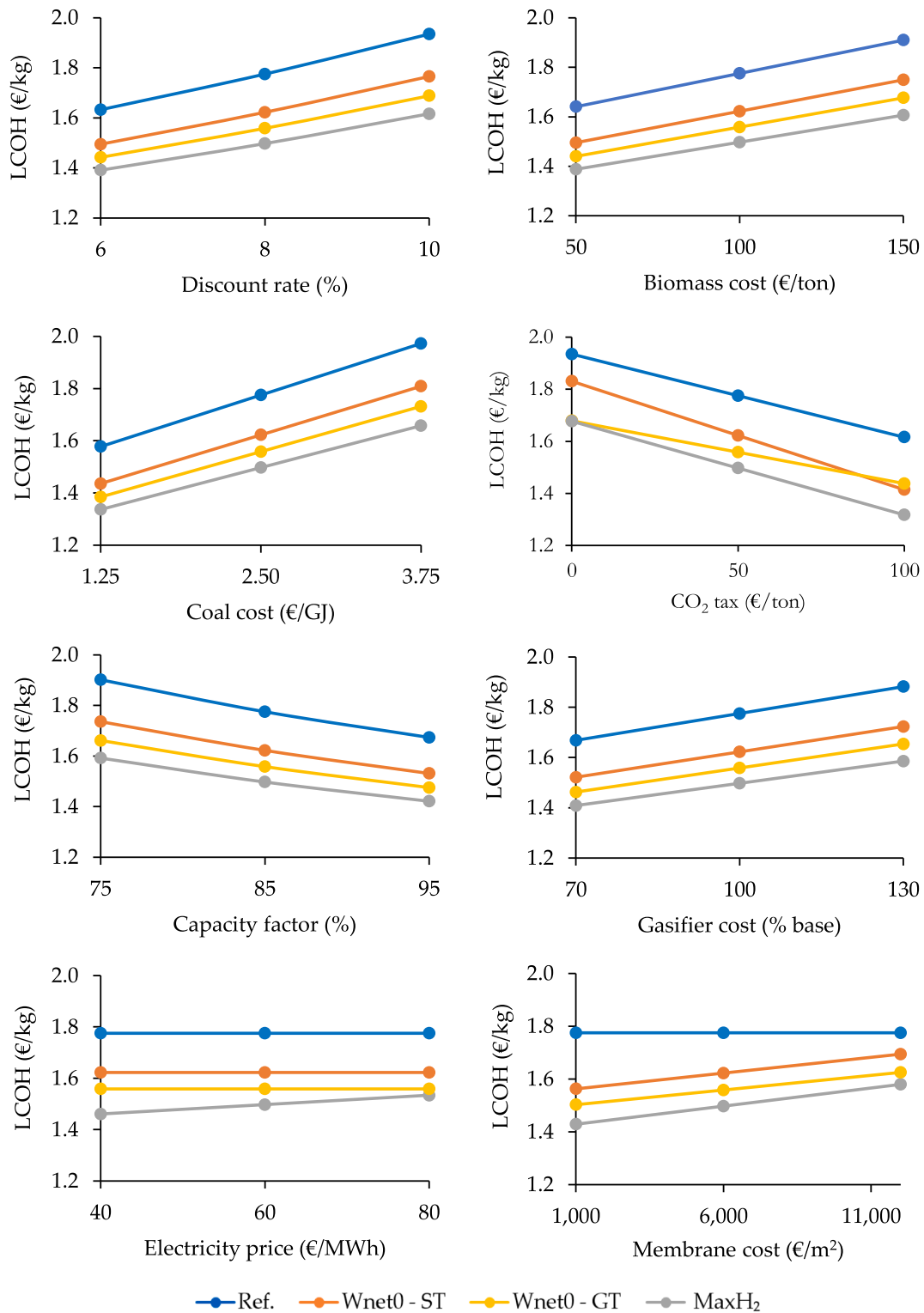


Fig. 7. Sensitivity analysis results.

### 3.2.3. Effect of biomass fraction

The extra cost of biomass fuel per unit of energy relative to coal can be outweighed by the increased revenue resulting from larger negative CO<sub>2</sub> emissions. For the biomass cost and CO<sub>2</sub> price levels considered in the base case models, the LCOH variation for the MaxH<sub>2</sub> model for biomass fractions ranging from 0 to 40%w. of the feed is shown in Fig. 8, with the relative contributions of each cost item. As shown, higher biomass fractions are economically beneficial under the base

assumptions, despite a mild reduction in H<sub>2</sub> production efficiency is due to decreasing CGE and increasing methane fraction (which cannot be converted to H<sub>2</sub> in the MAWGS reactor) when the biomass weight fraction increases. Electricity consumption slightly decreases for a larger biomass content, somewhat reducing the power imports and hence operating costs.

As the coal fraction reaches 100%, a smaller ASU is needed due to the decreased O<sub>2</sub> consumption in the gasifier. On the other hand, the O<sub>2</sub>

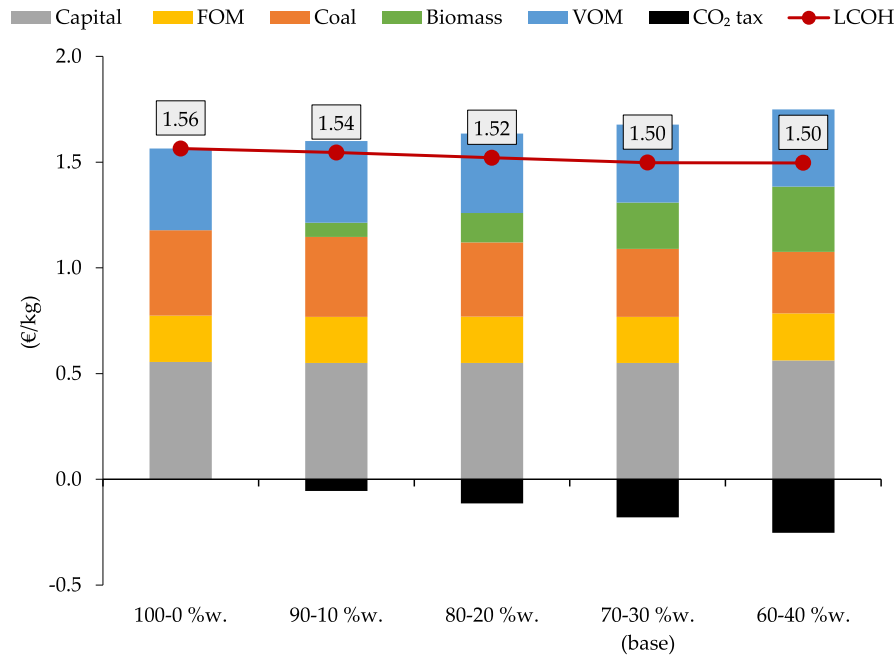


Fig. 8. LCOH of the MaxH<sub>2</sub> plant for different coal-biomass fractions in %w.

demand for the oxy-combustion of the retentate stream begins to increase as the biomass fraction rises due to a larger methane content in the syngas, slightly reducing H<sub>2</sub> efficiencies, which slightly reduces the economic benefit of higher biomass co-gasification fractions. However, these effects are small, and the main determinant of the economic attractiveness of higher biomass fractions is whether the CO<sub>2</sub> price cancels out the price premium of biomass. In Fig. 9, the break-even cost of biomass for different CO<sub>2</sub> prices is represented, as well as the base case assumptions (CO<sub>2</sub> price of 50 €/ton and biomass cost of 100 €/ton). It reflects the price region for biomass and CO<sub>2</sub> at which it is profitable to maximize the fraction of biomass intake to the plant, under the assumption that capital, FOM and other VOM costs remain constant.

If H<sub>2</sub> generated with this technology were to replace the European Union natural gas consumption of 2019 (around 17 EJ, based on the data from BP [48]), this would imply a yearly removal of approximately 0.6 gigatons of CO<sub>2</sub> from the atmosphere – fully 18% of total CO<sub>2</sub> emissions from 2018 [48]. Sufficient biomass from wood crops can be effectively supplied for this purpose based on several forecasts for Europe (de Wit et al. [49] Panoutsou et al. [50]).

### 3.2.4. Maximum economic potential

The previous assessments of advanced configurations accounted for

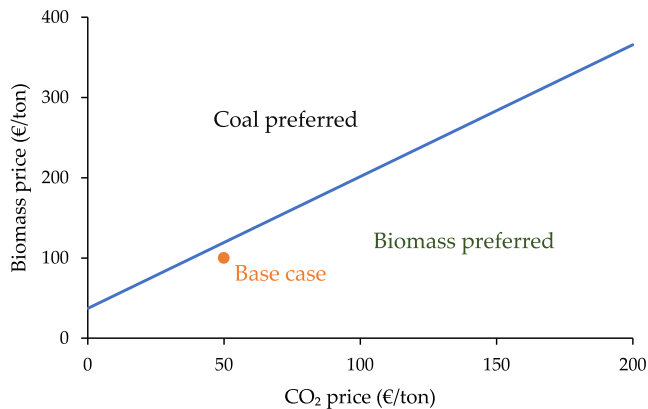


Fig. 9. Break-even biomass and CO<sub>2</sub> prices.

process contingency factors in those plant units which present a low technological maturity. In this section, to unravel the maximum economic potential of MaxH<sub>2</sub> case, an assessment considering no process contingency is presented. On top of that, the MaxH<sub>2</sub> model is modified to account for potential revenues from hot water production for district heating (DH) which can be produced from residual heat resulting in a combined heat and hydrogen generation plant (CHH<sub>2</sub>). Such residual heat is available in the condensing enthalpy of the permeate stream, roughly consisting of 50%mol of steam, which cannot be effectively retrieved for useful purposes within the original plant scheme and was inevitably rejected to the ambient. Given the permeate pressure of around 15 bar and the composition, condensation of the steam takes place in a wide temperature, starting at approximately 160 °C. This allows for an efficient heat exchange with pressurize water (16 bar) to around 120 °C (specifications for DH [51]), as illustrated in Fig. 10. The water from district heating at low temperature of 60 °C is pressurized again and sent to the heat recovery unit. 5% of the heat is assumed to be lost due to transmission. A large uncertainty relative to the selling price of hot water exists, and in this study a reasonable value of 30€/MWh is taken (including transportation costs).

Utilizing this remnant heat to generate hot water has a direct effect on the size of the cooling water tower and auxiliary consumption for heat rejection, presenting therefore a somewhat lower capital

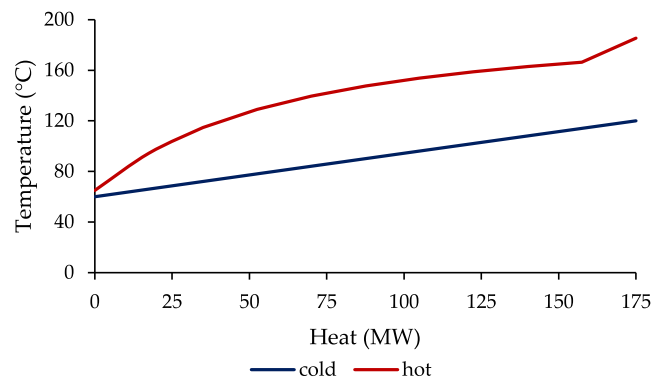


Fig. 10. Heat exchange profile in hot water generator.

investment for this item and smaller make-up water requirements relative to the original case. On the other hand, some extra power consumption is need for liquid pumping to overcome the pressure losses of the system. In Fig. 11, the LCOH breakdown for the cases discussed is presented. Avoiding process contingencies under the assumption that the more uncertain plant elements have reached a sufficient degree of technological maturity has a notable influence in the LCOH of the scheme, achieving a reduction of around 0.07 €/kg of H<sub>2</sub> (4.7% cheaper). Furthermore, when revenues from DH are accounted for, an additional cost reduction of around 0.19 €/kg of H<sub>2</sub> is achieved (17.3% overall reduction), which is even more significant, considering that the heat selling price is half of the electricity price assigned to the power imports for this plant.

#### 4. Summary and conclusions

A detailed techno-economic assessment of four negative emission technologies (NETs) for H<sub>2</sub> production from co-gasification of coal and biomass was carried out in the present study. The main results are summarized below:

- The reference plant employed a GE gasifier, Selexol selective H<sub>2</sub>S and CO<sub>2</sub> removal unit, PSA H<sub>2</sub> purification, and a steam power cycle, achieving a leveled cost of hydrogen (LCOH) of 1.78 €/kg and 93.8% CO<sub>2</sub> capture.
- Replacement of cold gas clean-up, Selexol absorption and PSA unit with hot gas clean-up (HGCU) and a membrane-assisted water-gas shift (MAWGS) reactor with subsequent oxycombustion of the retentate reduced the LCOH to 1.62 €/kg and achieved 100% CO<sub>2</sub> capture.
- Introduction of a more efficient E-gas gasifier with slurry vaporization to increase CGE, and power generation with a gas turbine fuelled by the purge stream of the CO<sub>2</sub> cryogenic purification unit (CPU) after the MAWGS, led to a LCOH to 1.56 €/kg while the CO<sub>2</sub> capture decreased to 91.7%.
- When the power cycle was removed, allowing for electricity imports and carrying out oxycombustion of the retentate to raise steam, hydrogen production could be maximized to reach a 1.50 €/kg LCOH achieving 100% CO<sub>2</sub> capture.

LCOH sensitivity to a wide range of economic assumptions was evaluated, with CO<sub>2</sub> price being the most influential parameter. All

plants benefitted strongly from higher CO<sub>2</sub> prices, particularly those which achieved 100% CO<sub>2</sub> capture. Under the base-case biomass and CO<sub>2</sub> prices of 100 and 50 €/ton, respectively, maximizing the biomass feed fraction was beneficial, decreasing the LCOH by 0.07 €/kg when the feed fraction increased from 0 to 40%w. Break-even biomass and CO<sub>2</sub> prices were determined, revealing a great potential for this bio-CCS technology to remove CO<sub>2</sub> from the atmosphere competitively by partially replacing coal as fuel. A final case to determine the maximum economic potential assumes no process contingencies for the less mature process units to reach a LCOH of 1.43 €/kg. Furthermore, when revenues from hot water production for district heating are accounted for (generated using the permeate stream condensation enthalpy), the LCOH reduces to 1.24 €/kg (~30% below the reference plant). Such a combined heat and hydrogen plant would be suitable to be deployed close to populated areas given the minimal emission levels attained through advanced gas clean-up systems, water quenching and NOx-free combustion in the oxygen carrier assisted combustor.

Key to the realization of this promising economic performance is the commercialization of high temperature syngas clean up, reliable operation of membrane technology, successful implementation of oxygen carrier-assisted combustion of the retentate, while achieving the solid fuel gasification enhancements (slurry vaporization, complete tar-heavy hydrocarbon conversion) that maximize cold gas efficiency, thereby limiting the partial oxidation losses and by-product formation. However, the relative simplicity of the MaxH<sub>2</sub> plant topology relative to the other cases suggests that this concept would be comparatively easier to build and to operate, provided that these technological hurdles are overcome. Further studies exploring the potential of increasing biomass feed fractions beyond the 30%-level assumed as the base-case are also recommended to unlock further cost reductions.

In conclusion, the NET plants evaluated in this study show great promise for providing a carbon-negative energy vector (H<sub>2</sub>) using abundant locally available solid fuels. Even the reference plant presents an economic performance that can be attractive in an environment with high CO<sub>2</sub> and moderate biomass prices. The advanced concepts further improve competitiveness relative to published works evaluating novel H<sub>2</sub> production technologies with CCS from natural gas feedstocks [4,7,9]. The cost-effective negative emissions from these plants can allow the hydrogen economy to play a leading role in achieving net-zero in a timely and economical manner.

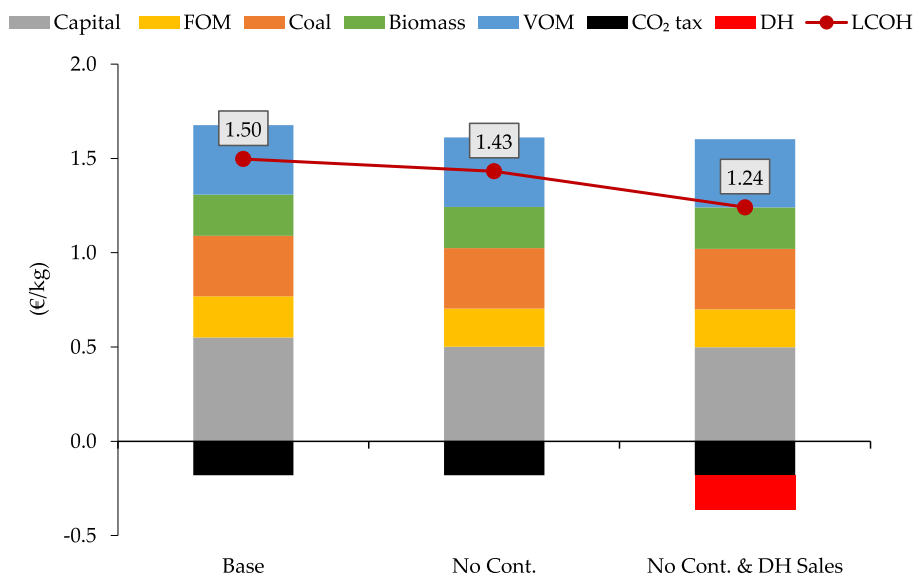


Fig. 11. LCOH for the MaxH<sub>2</sub> plant assuming no process contingency and revenues from DH.

## CRedit authorship contribution statement

**Carlos Arnaiz del Pozo:** Conceptualization, Methodology, Formal analysis, Investigation, Writing - original draft. **Schalk Cloete:** Conceptualization, Methodology, Formal analysis, Writing - review & editing, Funding acquisition. **Ángel Jiménez Álvaro:** Writing - review & editing, Supervision.

## Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

## Acknowledgement

The authors would like to acknowledge Honeywell for the free academic license of Unisim Design R451. The authors would also like to acknowledge AmsterCHEM for the free academic license of the CAPE-OPEN Scilab-Unisim unit operation. The authors from Universidad Politécnica de Madrid would like to acknowledge the Flow Technology Group from SINTEF Industry for the funding received to carry out this work within the framework of the project entitled “Techno-economic assessment of advanced H<sub>2</sub> and power plants”.

## Appendix A. Supplementary data

Supplementary data to this article can be found online at <https://doi.org/10.1016/j.enconman.2021.114712>.

## References

- [1] Anonymous. The Future of Hydrogen. 2019; IEA: Paris.
- [2] Cloete Schalk, Ruhnau Oliver, Hirth Lion. On capital utilization in the hydrogen economy: The quest to minimize idle capacity in renewables-rich energy systems. *Int J Hydrogen Energy* 2021;46(1):169–88.
- [3] Voldsund M, Jordal K, Anantharaman R. Hydrogen production with CO<sub>2</sub> capture. *Int J Hydrogen Energy*. 2016;41(9):4969–92.
- [4] Spallina V, Pandolfo D, Battistella A, Romano MC, Van Sint Annaland M, Gallucci F. Techno-economic assessment of membrane assisted fluidized bed reactors for pure H<sub>2</sub> production with CO<sub>2</sub> capture. *Energy Convers Manage* 2016; 120:257–73.
- [5] Khojasteh Salkuyeh Y, Saville BA, MacLean HL. Techno-economic analysis and life cycle assessment of hydrogen production from natural gas using current and emerging technologies. *Int J Hydrogen Energy* 2017;42(30):18894–909.
- [6] Spallina V, Marinello B, Gallucci F, Romano MC, Van Sint Annaland M. Chemical looping reforming in packed-bed reactors: modelling, experimental validation and large-scale reactor design. *Fuel Process Technol* 2017;156:156–70.
- [7] Cloete S, Khan MN, Amini S. Economic assessment of membrane-assisted autothermal reforming for cost effective hydrogen production with CO<sub>2</sub> capture. *Int J Hydrogen Energy*. 2019;44(7):3492–510.
- [8] Khan MN, Shamim T. Techno-economic assessment of a plant based on a three reactor chemical looping reforming system. *Int J Hydrogen Energy* 2016;41(48): 22677–88.
- [9] Nazir Shareq Mohd, Cloete Jan Hendrik, Cloete Schalk, Amini Shahriar. Pathways to low-cost clean hydrogen production with gas switching reforming. *Int J Hydrogen Energy* 2021;46(38):20142–58.
- [10] Yan Y, Manovic V, Anthony EJ, Clough PT. Techno-economic analysis of low-carbon hydrogen production by sorption enhanced steam methane reforming (SE-SMR) processes. *Energy Convers Manage* 2020;226:113530.
- [11] Lee H, Lee B, Byun M, Lim H. Comparative techno-economic analysis for steam methane reforming in a sorption-enhanced membrane reactor: Simultaneous H<sub>2</sub> production and CO<sub>2</sub> capture. *Chem Eng Res Design* 2021;171:383–94.
- [12] IEA. Hydrogen production costs using natural gas in selected regions. Paris: IEA; 2018.
- [13] IEAGHG CO. Capture at coal based power and hydrogen plants, Report 2014/3; 2014.
- [14] Basile A, Gallucci F, Paturzo L. A dense Pd/Ag membrane reactor for methanol steam reforming: experimental study. *Catal Today* 2005;104(2–4):244–50.
- [15] Uemiyama S, Sato N, Ando H, Kikuchi E. The water gas shift reaction assisted by a palladium membrane reactor. *Ind Eng Chem Res* 1991;30(3):585–9.
- [16] Jordal K, Anantharaman R, Peters TA, Berstad D, Morud J, Nekså P, et al. High-purity H<sub>2</sub> production with CO<sub>2</sub> capture based on coal gasification. *Energy* 2015; 88:9–17.
- [17] Campanari S, Mastropasqua L, Gazzani M, Chiesa P, Romano MC. Predicting the ultimate potential of natural gas SOFC power cycles with CO<sub>2</sub> capture – Part A: Methodology and reference cases. *J Power Sources* 2016;324:598–614.
- [18] Giuffrida A, Romano M, Lozza G. Thermodynamic Assessment of IGCC Plants with Hot Gas Desulphurization. *Appl Energy* 2010;87(11):3374–83.
- [19] Nexant I. Preliminary feasibility analysis of RTI warm gas cleanup (WGCU) technology; 2007.
- [20] Higman C. Gasification. 2nd ed.; 2008.
- [21] Sikarwar VS, Zhao M, Clough P, Yao J, Zhong X, Memon MZ, et al. An overview of advances in biomass gasification. *Energy Environ Sci* 2016;9(10):2939–77.
- [22] Fajardy M, Mac Dowell N. Can BECCS deliver sustainable and resource efficient negative emissions? *Energy Environ Sci* 2017;10(6):1389–426.
- [23] Hetland J, Yowargana P, Leduc S, Kraxner F. Carbon-negative emissions: Systemic impacts of biomass conversion: a case study on CO<sub>2</sub> capture and storage options. *Int J Greenhouse Gas Control* 2016;49:330–42.
- [24] Kapetaki Z, Brandani S, Brandani P, Ahn H. Process simulation of a dual-stage Selexol process for 95% carbon capture efficiency at an integrated gasification combined cycle power plant. *Int J Greenhouse Gas Control* 2015;39:17–26.
- [25] Jones D, Bhattacharyya D, Turton R, Zitney SE. Optimal design and integration of an air separation unit (ASU) for an integrated gasification combined cycle (IGCC) power plant with CO<sub>2</sub> capture. *Fuel Process Technol* 2011;92(9):1685–95.
- [26] Qin K, Lin W, Jensen PA, Jensen AD. High-temperature entrained flow gasification of biomass. *Fuel* 2012;93:589–600.
- [27] Anantharaman R, Bolland O, Booth N, Van Dorst E, Sanchez Fernandez E, Franco F, et al. Cesar Deliverable D2.4.3. European Best Practice Guidelines For Assessment Of Co<sub>2</sub> Capture Technologies; 2018.
- [28] Quian Zhu. High Temperature Syngas Coolers, CCC/257. IEA Clean Coal Centre; 2015.
- [29] Nazir SM, Cloete JH, Cloete S, Amini S. Efficient hydrogen production with CO<sub>2</sub> capture using gas switching reforming. *Energy* 2019;185:372–85.
- [30] Gräbner M, Meyer B. Performance and exergy analysis of the current developments in coal gasification technology. *Fuel* 2014;116:910–20.
- [31] Aiuchi K, Moriyama R, Takeda S, Kitada S, Onozaki M, Katayama Y. A pre-heating vaporization technology of coal-water-slurry for the gasification process. *Fuel Process Technol* 2007;88(4):325–31.
- [32] <https://gasforclimate2050.eu/ehb/> 2021.
- [33] Kolster C, Mechleri E, Krevor S, Mac Dowell N. The role of CO<sub>2</sub> purification and transport networks in carbon capture and storage cost reduction. *Int J Greenhouse Gas Control* 2017;58:127–41.
- [34] Szima S, Arnaiz del Pozo C, Cloete S, Chiesa P, Jiménez Alvaro Á, Cormos A, et al. Finding synergy between renewables and coal: Flexible power and hydrogen production from advanced IGCC plants with integrated CO<sub>2</sub> capture. *Energy Convers Manage* 2021;231:113866.
- [35] Carlos Arnaiz del Pozo, Schalk Cloete, Ángel Jiménez Álvaro. Standard Economic Assessment (SEA) Tool. Available from: <https://bit.ly/3hyFITT>.
- [36] Carlos Arnaiz del Pozo, Schalk Cloete and Ángel Jiménez Álvaro. SEA Tool User Guide. Available from: <https://bit.ly/3jq9Bkf>.
- [37] Carlos Arnaiz del Pozo, Schalk Cloete, Ángel Jiménez Álvaro. Carbon Negative Hydrogen - BioCCS Economic evaluations. Available from: <https://bit.ly/2ULkd9k>.
- [38] Turton R, Bailie RC, Whiting WB, Shaeiwitz JA. Analysis, synthesis and design of chemical processes; 2008.
- [39] IEAGHG R&D Programme. Effects of Plant Location on the Costs of CO<sub>2</sub> Capture, Technical Report; 2018.
- [40] Plant Cost Index Archives – Chemical Engineering. <http://www.chemengonline.com> 2018.
- [41] R.2. DOE/NETL-2010/1397. Cost performance baseline for fossil energy plants. Volume 1: Bituminous coal and natural gas to electricity. Performance baseline for fossil energy plants. Volume 1: Bituminous coal and natural gas to electricity; 2013.
- [42] Zoelle AJ, Turner MJ, Woods MC, James III PhD, Robert E, Fout TE, Shultz TR. Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 4. In Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 4. 2018.
- [43] Roussanaly S, Rubin E, Der Spek MV, Berghout N, Booras G, Fout T, Garcia M, Gardarsdottir S, Matuszewski M, McCoy S. Towards improved guidelines for cost evaluation of CO<sub>2</sub> capture technologies; 2019.
- [44] Hirth L, Steckel JC. The role of capital costs in decarbonizing the electricity sector. *Environ Res Lett* 2016;11(11):114010.
- [45] Nazir SM, Morgado JF, Bolland O, Quinta-Ferreira R, Amini S. Techno-economic assessment of chemical looping reforming of natural gas for hydrogen production and power generation with integrated CO<sub>2</sub> capture. *Int J Greenhouse Gas Control* 2018;78:7–20.
- [46] Rosner F, Chen Q, Rao A, Samuelsen S, Jayaraman A, Alptekin G. Process and economic data for the thermo-economic analyses of IGCC power plants employing warm gas CO<sub>2</sub> separation technology. *Data Brief* 2019;27:104716.
- [47] Arratibel Plazaola A, Pacheco Tanaka DA, Van Sint Annaland M, Gallucci F. Recent advances in Pd-based membranes for membrane reactors. *Molecules* 2017;22(1): 51.
- [48] Anonymous “<https://www.bp.com/en/global/corporate/energy-economics/statistical-review-of-world-energy.html>” 2021.

- [49] de Wit M, Faaij A. European biomass resource potential and costs. *Biomass Bioenergy* 2010;34(2):188–202.
- [50] Panoutsou C, Eleftheriadis J, Nikolaou A. Biomass supply in EU27 from 2010 to 2030. *Energy Policy* 2009;37(12):5675–86.
- [51] Anonymous. [https://www.euroheat.org/wp-content/uploads/2016/04/UP-RES\\_M6\\_District\\_Heating\\_and\\_Cooling.pdf](https://www.euroheat.org/wp-content/uploads/2016/04/UP-RES_M6_District_Heating_and_Cooling.pdf); 2021.