



## Research Paper

# Synthetic natural gas as a green hydrogen carrier – Technical, economic and environmental assessment of several supply chain concepts

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## ARTICLE INFO

## Keywords:

CO<sub>2</sub> cycle  
Green carbon  
Green hydrogen  
Hydrogen supply  
Hydrogen supply chain assessment  
Synthetic natural gas

## ABSTRACT

Based on synthetic natural gas, existing natural gas markets and infrastructures can be used to make renewable sources of energy from sun- and/or wind-rich regions available on a global scale. To overcome the challenge of providing non-fossil CO<sub>2</sub> for the production of this synthetic natural gas, a novel concept analyzed in this paper envisages to reform the synthetic natural gas in the importing country and transporting the captured CO<sub>2</sub> back to the exporting country to be reused for the production of synthetic natural gas; i.e., the synthetic natural gas serves as a hydrogen carrier. This paper examines and compares the energy efficiency, cost and greenhouse gas emissions of different hydrogen supply chains using synthetic natural gas as a carrier related to the year 2030. To do so, all relevant components are taken into account to model the entire supply chains. A special focus is put on different options for providing the required CO<sub>2</sub> and on different technologies for synthetic natural gas reforming. The assessment shows that the availability of a cheap source of biogenic CO<sub>2</sub> at the point-of-export as well as electrified steam methane reforming result in the lowest hydrogen supply cost of 6.6 to 7.0 €<sub>2020</sub>/kgH<sub>2</sub>, also achieving best results in terms of energy efficiency (around 44%). With regard to minimizing greenhouse gas emissions, autothermal reforming of methane appears to be advantageous. A closed CO<sub>2</sub> cycle is favorable over sole onsite CO<sub>2</sub> provision, if no cheap CO<sub>2</sub> of non-fossil origin is available at the point-of-export and the costly direct air capture process would have to be used.

## 1. Introduction and background

International energy trading is crucial to cover the energy demand of many countries. In recent years, especially the maritime transportation of liquefied natural gas (LNG) has steadily gained significance. Thus, the volume of LNG traded worldwide almost doubled between 2010 and 2023 [1,2]. This LNG market is expected to continue to grow unabated in the coming years, particularly in light of Europe's abandonment of pipeline gas from Russia, and the sustained growth in energy demand in the People's Republic of China, one of the largest importers of LNG [3]. The dynamic growth of the LNG trade has necessitated a strong expansion of the corresponding infrastructure. For example, based on regasification terminals recently completed and currently under development, the available LNG import capacity in the European Union (EU) is expected to roughly double between 2020 and 2030 [4–6].

In parallel to these developments, the pressure to supply society with climate-neutral energy is increasing due to the obligations agreed in the Paris Climate Agreement and the more and more apparent negative and harmful consequences of climate change [7,8]. In this context, the

discussion about the use of the LNG import infrastructure within a fossil-free future is being pursued with vigor, particularly within the European Union (EU). Concepts in which the LNG regasification terminals are converted to import liquefied “green” hydrogen or renewable ammonia are being discussed more frequently [9–12]. A technically simple alternative for the continued utilization of the global LNG transportation infrastructure is the development of a market for renewable synthetic natural gas (SNG) that is chemically identical to natural gas of fossil origin. Thus, SNG not only offers the opportunity to transfer existing infrastructures and markets to a fossil-free future energy system, but could also help to supply countries such as Germany, Japan or South Korea, which will likely remain dependent on energy imports for decades to come, with renewable energy [13,14].

In order to provide SNG on a systemically relevant scale, large quantities of carbon dioxide (CO<sub>2</sub>) are required in addition to renewable (“green”) hydrogen. Within an energy system that is in line with the Paris Agreement and does not release any additional greenhouse gases (GHG) into the atmosphere, the CO<sub>2</sub> used for SNG production must be of non-fossil origin, assuming the CO<sub>2</sub> is not stored underground but emitted into the atmosphere during SNG use. Due to the currently

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## Nomenclature

### Abbreviation Meaning

$ACAPEX_i$	Annuity of capital expenditures of supply chain component $i$ , €/2020/a	$i$	$g_{CO_2\text{-eq}}$ Components of the synthetic natural gas supply chain including the components of the carbon dioxide cycle
$AF_i$	Annuity factor of supply chain component $i$ , -	$INFL$	Projected future inflation rate, -
ATR	Autothermal reforming	KOH	Potassium hydroxide
Biog.	Biogenic carbon dioxide source	LCO <sub>2</sub>	Liquefied carbon dioxide
$c_{CO_2\text{ liquefaction}}$	Cost for carbon dioxide liquefaction, €/2020/kg <sub>H2</sub>	LHV	Lower heating value
$c_{CO_2\text{ supply}}$	Total cost for continuous carbon dioxide supply at the point-of-export, €/2020/kg <sub>H2</sub>	LNG	Liquefied natural gas
$c_{CO_2\text{ storage}}$	Cost for carbon dioxide storage, €/2020/kg <sub>H2</sub>	LPG	Liquefied petroleum gas
$c_{CO_2\text{ supply, onsite}}$	Cost for onsite carbon dioxide supply via direct air capture or biogenic source, €/2020/kg <sub>H2</sub>	LSNG	Liquefied synthetic natural gas
$c_{CO_2\text{ transport}}$	Cost for carbon dioxide transportation via ship, €/2020/kg <sub>H2</sub>	$LT_i$	Technical lifetime supply chain component $i$ , a
$c_{\text{component},i}$	Specific cost of the supply chain component $i$ , €/2020/kg <sub>H2</sub>	$\dot{m}_{H_2\text{ at IP}}$	Amount of hydrogen provided at the point-of-import, kg <sub>H2</sub> /a
$c_{\text{conversion}}$	Cost for methanation and synthetic natural gas liquefaction, €/2020/kg <sub>H2</sub>	NaOH	Sodium hydroxide
$c_{H_2\text{ at EP}}$	Cost for continuous “green” hydrogen supply at the point-of-export, €/2020/kg <sub>H2</sub>	NG	Natural gas
$c_{H_2\text{ at IP}}$	Overall cost of hydrogen supplied at the point-of-import, €/2020/kg <sub>H2</sub>	O <sub>2</sub>	Oxygen
$c_{\text{reconversion}}$	Cost for reconversion of synthetic natural gas into compressed, purified hydrogen, €/2020/kg <sub>H2</sub>	OPEXi	Operational expenditures of supply chain component $i$ , €/2020/a
$c_{\text{storage}}$	Cost for synthetic natural gas storage, €/2020/kg <sub>H2</sub>	PEM	Polymer electrolyte membrane
$c_{\text{transport}}$	Cost for synthetic natural gas transportation via ship, €/2020/kg <sub>H2</sub>	POX	Partial oxidation
CaCO <sub>3</sub>	Calcium carbonate	PSA	Pressure swing adsorption
$CAPEX_i$	Capital expenditure of supply chain component $i$ , €/2020	$Q_{CO_2\text{ liquefaction}}$	Energy input for carbon dioxide liquefaction, kWh/kg <sub>H2</sub>
CO	Carbon monoxide	$Q_{CO_2\text{ supply}}$	Total energy input for continuous carbon dioxide supply at the point-of-export, kWh/kg <sub>H2</sub>
CO <sub>2</sub>	Carbon dioxide	$Q_{CO_2\text{ supply, cycle}}$	Energy input for the return of carbon dioxide from the point-of-import to the point-of-export, kWh/kg <sub>H2</sub>
CO <sub>2-eq</sub>	Carbon dioxide equivalent	$Q_{CO_2\text{ supply, onsite}}$	Energy input for onsite carbon dioxide supply via direct air capture or biogenic source, kWh/kg <sub>H2</sub>
DAC	Direct air capture	$Q_{CO_2\text{ storage}}$	Energy input for carbon dioxide storage, kWh/kg <sub>H2</sub>
$di$	Depreciation time, a	$Q_{CO_2\text{ transport}}$	Energy input for carbon dioxide transportation via ship, kWh/kg <sub>H2</sub>
$e_{CO_2\text{ supply}}$	Total emissions of continuous carbon dioxide supply at the point-of-export, g <sub>CO<sub>2-eq</sub></sub> /kg <sub>H2</sub>	$Q_{\text{conversion}}$	Energy input for methanation and synthetic natural gas liquefaction, kWh/kg <sub>H2</sub>
$e_{\text{component},i}$	Emissions of supply chain component $i$ , g <sub>CO<sub>2-eq</sub></sub> /kg <sub>H2</sub>	$Q_{\text{conversion loss}}$	Energy losses during conversion from hydrogen to liquefied synthetic natural gas, kWh/kg <sub>H2</sub>
$e_{\text{conversion}}$	Emissions from methanation and synthetic natural gas liquefaction, g <sub>CO<sub>2-eq</sub></sub> /kg <sub>H2</sub>	$Q_{H_2\text{ at EP}}$	Energy input for continuous “green” hydrogen supply at the point-of-export, kWh/kg <sub>H2</sub>
$e_{\text{energy},H_2\text{ at EP}}$	Emissions of energy use for hydrogen supply at the point-of-export, g <sub>CO<sub>2-eq</sub></sub> /kg <sub>H2</sub>	$Q_{H_2\text{ at IP}}$	Energy content of the hydrogen supplied at the point-of-import, kWh/kg <sub>H2</sub>
$e_{\text{energy},i}$	Emissions of supply chain component $i$ caused by the use of energy, g <sub>CO<sub>2-eq</sub></sub> /kg <sub>H2</sub>	$Q_{H_2\text{ at EP, ideal}}$	Energy input at point-of-export for the supply of hydrogen effectively delivered at the point-of-import, kWh/kg <sub>H2</sub>
$e_{\text{fac,elec}}$	Emission factor of a renewable electricity supply to the hydrogen production, g <sub>CO<sub>2-eq</sub></sub> /kW <sub>hel</sub>	$Q_{H_2\text{ SC losses}}$	Energy input at the point-of-export to compensate losses along the supply chain, kWh/kg <sub>H2</sub>
$e_{H_2\text{ at EP}}$	Emissions from continuous “green” hydrogen supply at the point-of-export, g <sub>CO<sub>2-eq</sub></sub> /kg <sub>H2</sub>	$Q_{\text{input}}$	Total energy input along all supply chain components, kWh/kg <sub>H2</sub>
$e_{H_2\text{ at IP}}$	Overall emissions of hydrogen supplied at the point-of-import, g <sub>CO<sub>2-eq</sub></sub> /kg <sub>H2</sub>	$Q_{\text{reconversion}}$	Energy input for reconversion of synthetic natural gas into compressed, purified hydrogen, kWh/kg <sub>H2</sub>
$e_{\text{leakage},i}$	Emissions of supply chain component $i$ caused by leakage of greenhouse gases, g <sub>CO<sub>2-eq</sub></sub> /kg <sub>H2</sub>	$Q_{\text{reconversion loss}}$	Energy losses during reconversion of synthetic natural gas into compressed, purified hydrogen, kWh/kg <sub>H2</sub>
$e_{\text{reconversion}}$	Emissions from reconversion of synthetic natural gas into compressed, purified hydrogen, g <sub>CO<sub>2-eq</sub></sub> /kg <sub>H2</sub>	$Q_{\text{storage}}$	Energy input for synthetic natural gas storage, kWh/kg <sub>H2</sub>
$e_{\text{storage}}$	Emissions from liquefied synthetic natural gas storage, g <sub>CO<sub>2-eq</sub></sub> /kg <sub>H2</sub>	$Q_{\text{storage loss}}$	Energy losses during synthetic natural gas storage, kWh/kg <sub>H2</sub>
$e_{\text{transport}}$	Emissions from synthetic natural gas transportation via ship, g <sub>CO<sub>2eq</sub></sub> /kg <sub>H2</sub>	$Q_{\text{transport}}$	Energy input for synthetic natural gas transportation via ship, kWh/kg <sub>H2</sub>
el.	Electricity	$Q_{\text{transport loss}}$	Energy losses during synthetic natural gas transportation via ship, kWh/kg <sub>H2</sub>
eSMR	Electrified steam methane reforming	RWGS	Reverse water-gas shift reaction
EU	European Union	SMR	Steam methane reforming
GHG	Greenhouse gas	$WACC_{\text{nom}}$	Nominal weighted average cost of capital, -
GH <sub>2</sub>	Gaseous hydrogen	$WACC_{\text{real}}$	Real weighted average cost of capital, -
GWP <sub>100</sub>	Global warming potential for a time period over 100 years,	WGS	Water-gas shift reaction
		$\eta_{\text{SC}}$	Overall point-to-point hydrogen supply chain energy efficiency, %

limited and uncertain future availability of CO<sub>2</sub> of non-fossil origin, the implementation of technically closed CO<sub>2</sub> cycles is an innovative option to enable “green” energy supply chains based on SNG and other synthetic hydrocarbons [15,16].

Within such a concept, the CO<sub>2</sub> produced during SNG utilization would be captured and subsequently returned to the site of the “green” hydrogen production to be available there again for the generation of SNG. To enable a closed CO<sub>2</sub> cycle with a high overall CO<sub>2</sub> cycling rate, the SNG would most likely have to be used at the place of destination in large-scale units where an efficient CO<sub>2</sub> capturing is possible. Options for large-scale SNG application might be the use in gas-fired power plants as well as in reforming units for the provision of hydrogen. The hydrogen produced in this way could thereafter be used flexibly in various applications within the overall economy. Such a SNG-based hydrogen supply competes with other “green” hydrogen supply options (e.g., liquid hydrogen, ammonia, methanol) [17–19].

Against this background, this paper aims to provide a better understanding of SNG-based hydrogen supply while assessing novel approaches for key components of this supply chain such as the provision of CO<sub>2</sub> and methane reforming. Especially the CO<sub>2</sub> supply via a closed CO<sub>2</sub> cycle, as investigated in this study, represents an innovative approach that has been insufficiently addressed in the existing literature. At first, possible components of such SNG-based hydrogen supply chains and their current state of development are outlined. On this basis, potential configurations of SNG-based hydrogen supply chains are identified. Lastly, the identified supply chains are modeled and assessed related to the energy efficiency, cost and GHG emissions.

## 2. Hydrogen supply via SNG

Under ambient conditions, gaseous hydrogen is characterized by a very low volumetric energy density. Therefore, transportation over longer distances typically requires hydrogen conditioning to substantially enhance its volumetric energy density. This conditioning can consist of compressing or liquefying hydrogen. Here, hydrogen remains in its elementary state. In contrast, alternative options exist where hydrogen is either bound to a carrier molecule (e.g., to *Liquid Organic Hydrogen Carrier* – LOHC) or converted with another hydrogen-free molecule into a hydrogen-rich molecule (e.g., ammonia, methanol, SNG). To enable such hydrogen transportation in chemically bound forms, the respective carrier molecule must be provided at the point-of-export in addition to the hydrogen. After transportation, reconversion is necessary to supply pure hydrogen again. If hydrogen-rich molecules / hydrogen derivatives are employed for transportation, direct use may also be an option, depending on the intended end use. Fig. 1 illustrates the basic concept of supply chains for international hydrogen transportation and the above-mentioned key technical components. To allow for a fair comparison of these transportation options, this paper focuses

solely on the supply of pure hydrogen, neglecting the possibility of supplying hydrogen derivatives to the end consumer.

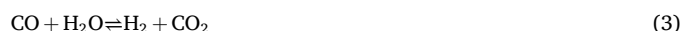
In order to create a holistic understanding of SNG-based hydrogen supply chains, the most important system components are discussed below. In addition to an overview of the fundamental technological principles, this section also provides an insight into the current state-of-the-art of the respective technology and the potential use of existing infrastructure components.

### 2.1. Conversion of hydrogen to SNG

Catalytic methanation is the conversion of hydrogen to methane. The methanation reaction, also known as the Sabatier process, can be realized based on both carbon monoxide (CO) and CO<sub>2</sub> (equation (1) and (2)) [20]. This reaction is typically controlled by suitable catalysts (e.g., nickel, ruthenium, cobalt) [21,22].



The two methanation reactions are interconnected through the water–gas shift reaction (WGS) (equation (3)) [20].



Commonly, CO<sub>2</sub> methanation is assumed to occur in a two-step process involving the reverse water–gas shift reaction (RWGS) followed by subsequent methanation of the generated CO [23,24]. The methanation reaction is influenced in particular by temperature and pressure. For instance, within the temperature range of 250 to 550 °C, where catalytic methanation typically occurs, maintaining pressures of at least 10 bar or higher is necessary to achieve high conversion rates [25].

The methanation reaction is strongly exothermic; i.e., part of the energy bound in the educt (hydrogen) is released as heat [26]. The exploitable temperature level of this waste heat depends on the respective process design, but is likely to be at least 150 °C in large-scale systems [27]. The maximum chemical conversion efficiency describing the transfer of the energy bound in the hydrogen to the methane is 83.2 % assuming a full conversion according to equation (1) and (2). In practical applications the theoretical maximum conversion efficiency can be nearly achieved [23,25,28].

Methanation using a CO-rich synthesis gas is an established industrial process that has been employed for many decades. One example are coal gasification units, where the largest plants reach capacities within the GW scale. In contrast, methanation with CO<sub>2</sub> being of particular relevance for SNG-based supply of “green” hydrogen, has so far only been proven on a pilot and demonstration scale [27].

The application of this technology outside of pure research environment has been demonstrated by a plant in Werlte, Germany,

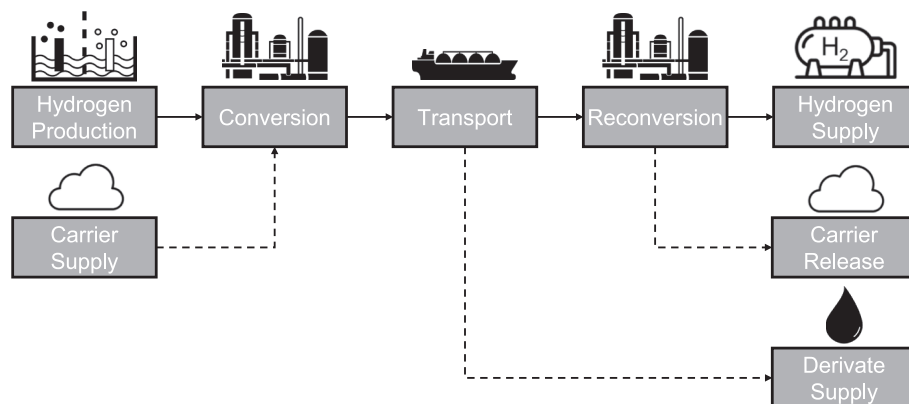


Fig. 1. Basic setup of hydrogen supply chains.

operating with a capacity of around 3.2 MW in terms of SNG output ( $\sim 230 \text{ kg}_{\text{SNG}}/\text{h}$ ) [27]. Consequently, a significant upscaling of catalytic methanation would be necessary for the large-scale implementation of SNG-based international energy supply chains.

## 2.2. Long-distance transportation of SNG

SNG transportation over distances of several thousand kilometers can be realized via gas pipelines and LNG carriers, as is the current practice in natural gas trade. For a future international trade of hydrogen and its derivatives, especially Europe and Asia (e.g., Japan and Korea) are expected to become important importers of “green” energy [13,14]. To interconnect these import markets with regions frequently discussed as exporters of “green” energy (e.g., Arabian Peninsula, Australia, North America, Namibia), pipeline connections are often no alternative due to the distance to be covered as well as geographical or political barriers. Therefore, only ship-based transportation via liquefied synthetic natural gas (LSNG) carriers is considered here.

To enable efficient shipping, SNG is liquefied to increase the energy density by a factor of 600. The technology necessary for such a methane liquefaction is identical to the units currently applied in plants for the liquefaction of fossil natural gas. Prior to liquefaction, the SNG must be purified (i.e., all unwanted components such as water,  $\text{CO}_2$  and nitrogen have to be removed from the synthetic methane) [29].

The liquefaction process involves a series of compression steps, heat exchanges and expansion steps to achieve a temperature below the boiling point of methane being  $-161.5 \text{ }^\circ\text{C}$  at atmospheric pressure. The SNG liquefaction process makes use of the Joule-Thomson effect of real gases according to which an expansion is accompanied by a cooling of the gas [29,30]. Energy, mainly in the form of electricity, is needed to operate such a liquefaction plant. Available data on the respective energy demand vary widely and range from  $0.25 \text{ kWh/kg}_{\text{SNG}}$  [31] to  $1.39 \text{ kWh/kg}_{\text{SNG}}$  [32].

LNG carriers are specifically designed vessels equipped with insulated cargo tanks to store LNG at cryogenic temperatures and keep it in liquid form. Due to the chemical identity, currently operated LNG carriers can also be used for the transportation of LSNG. LNG carriers under operation are usually equipped with membrane tanks or spherical tanks. Due to the better use of space and the lower weight, membrane tanks gained more and more market importance in recent years [33,34]. Especially very large LNG carriers, reaching transportation capacities of up to  $266,000 \text{ m}^3$ , are mainly equipped with membrane tanks [19,29].

During transportation, energy from the environment inevitably enters the LNG tanks, causing some of the (S)LNG to evaporate. For large LNG carriers, the boil-off rate is between 0.08 and 0.15 % per day of the transported gas. Smaller LNG carriers have typically higher boil-off rates due to the less favorable surface-to-volume ratio. Usually, the boil-off gas is used to power the ship, but re-liquefaction is also conceivable [33,35,36].

Today, natural gas is almost exclusively demanded in gaseous form. Hence, LNG is typically regasified directly at the import terminal. Also, for a hydrogen supply using SNG as a carrier, regasification at the import terminal appears to be the most suitable solution. In regasification terminals, LNG is commonly heated by seawater. The cold energy released within this process is currently not utilized in most cases. However, utilization of this cold is basically possible and could improve the overall energy efficiency of the process (e.g., electricity production in dedicated cryogenic power generation cycles, separation of gases, seawater desalination, liquefaction of other gas streams) [37].

Liquefaction plants, LNG carriers and regasification terminals have been used for decades for the intercontinental transportation of fossil LNG and are therefore available as proven technologies on a commercial scale. The future utilization of the existing LNG infrastructure for SNG supply is possible, provided that it is already available in the SNG exporting resp. importing country. However, this is not currently the case for a significant number of countries with promising potential for

the production of “green” hydrogen, and thus potentially also SNG [4,32].

## 2.3. Reconversion of SNG to hydrogen

For hydrogen production on the basis of natural gas, steam methane reforming (SMR) is today’s most commonly used process. SMR is a catalytically controlled process in which methane reacts with water vapor in a highly endothermic reaction to CO and hydrogen (equation (4)). Subsequently, a WGS is applied to oxidize the remaining CO with additional steam to  $\text{CO}_2$  and increase the hydrogen yield (equation (5)) [38,39]. These reactions take place at temperatures between  $650$  and  $1,000 \text{ }^\circ\text{C}$  and a pressure of  $5$  to  $40 \text{ bar}$  [40]. Further steps include hydrogen purification (e.g., pressure swing adsorption (PSA)).



To provide the thermal energy required for the reaction, heating of the reactor is necessary. Conventionally, SMR reactors are heated by burning additional natural gas. Typically, 30 to 40 % of the total natural gas input is used as heating fuel [41]. For SNG-based hydrogen supply involving a closed  $\text{CO}_2$  cycle, the  $\text{CO}_2$  created during methane reforming needs to be captured [39]. To achieve high  $\text{CO}_2$  capture rates, the  $\text{CO}_2$  must be captured from the syngas at the reformer outlet as well as from the flue gas of the gas burner heating the reformer. In the burner flue gas from natural gas combustion,  $\text{CO}_2$  is found in relatively low concentrations (5 to 10 vol-%) and at low pressure, implying a high capture effort [42,43]. Accordingly, for conventional SMR equipped with  $\text{CO}_2$  capture, which has not yet been realized on a large scale,  $\text{CO}_2$  capture rates of maximum 90 % are reported [42,44–46].

To overcome the technical challenges associated with  $\text{CO}_2$  capture in conventional SMR, various alternative processes are being discussed for methane reforming. Processes in which the  $\text{CO}_2$  does not occur diluted but highly concentrated at a single point are particularly promising.

One widely discussed alternative for methane-based hydrogen production meeting these requirements is autothermal reforming (ATR) [45,47–49]. Here, the endothermic SMR (equation (4)) is combined with exothermic partial oxidation (POX) (equation (6)) to enable the reformation to be operated adiabatically (heat-tight); i.e., no heat needs to be provided externally [47]. Accordingly,  $\text{CO}_2$  is produced exclusively in the reactor and can be captured comparatively easy. However, oxygen ( $\text{O}_2$ ) must be supplied to the reaction in order to enable the POX (equation (6)) [48].



Existing large-scale plants applying ATR technology are typically designed to provide CO- and hydrogen-rich synthesis gases. To date, there is no large-scale experience with hydrogen production by means of ATR equipped with  $\text{CO}_2$  capture [48]. When ATR is to be applied in the future for the production of pure hydrogen, e.g., as part of an SNG-based hydrogen supply chain, the resulting CO can be converted with water vapor within a WGS (equation (5)) to  $\text{CO}_2$  and hydrogen, increasing the hydrogen yield. The separation and purification of the product gases (hydrogen and  $\text{CO}_2$ ) can be realized by a combination of amine scrubbing and PSA [45,48]. Since all generated  $\text{CO}_2$  occur in a high concentration stream and have to be separated from the hydrogen in the purification unit anyway, high capture rates can be achieved with, compared to conventional SMR, less effort. Thus,  $\text{CO}_2$  capture rates of 95 % and more seem to be feasible for ATR [32,39,47,50,51]. The chemical conversion efficiency of the ATR – i.e. the share of energy bound in methane converted into energy bound in hydrogen – is between 75 and 80 % [45]. This corresponds to  $3.2$  to  $3.0 \text{ kg}_{\text{CH}_4}/\text{kg}_{\text{H}_2}$ . If the overall energy balance of the process is considered, the energy

demand upstream and downstream of the actual reforming must also be taken into account. In this context, the supply of pure O<sub>2</sub>, e.g., via an air separation unit, is of particular relevance. Furthermore, energy is also required for the separation and purification of the product gases [45].

Another approach to realize high CO<sub>2</sub> capture rates in hydrogen production from methane involves a modification of conventional SMR. Thus, electrically heated SMR reactors are currently being developed (electrified steam methane reforming – eSMR); a small-scale demonstration plant in Foulum, Denmark, is already under operation [52,53]. In such an eSMR reactor, the use of methane to meet the heating demand of the endothermic conversion reaction is fully omitted and the CO<sub>2</sub> is generated exclusively within the reforming reactor. Accordingly, CO<sub>2</sub> capture could be realized similar to ATR and focus on the separation from the synthesis gas at the reactor outlet [44]. CO<sub>2</sub> capture rates of 95 % and more are predicted to be feasible for eSMR at limited additional energy expense [39,44,54]. In addition, the deployment of the eSMR technology promises further advantages. For example, in conventional SMR process, the limited thermal conductivity between the catalyst and the reactor walls and the resulting steep temperature gradients lead to a poor utilization of the catalyst, resulting in less than 5 % of the reactor volume being used as an active zone for the conversion [55–57]. The eSMR applies electrical resistance heating direct to the catalyst and thereby overcomes these limitations leading to a significantly higher catalyst utilization. Consequently, eSMR reactors can potentially have a much smaller volume at the same hydrogen output and be operated more flexibly compared to conventional SMR reactors [55,57]. A key characteristic of the eSMR is the high potential hydrogen yield. The methane demand is estimated to be between 2.4 and 2.6 kg<sub>CH<sub>4</sub></sub>/kg<sub>H<sub>2</sub></sub> [39]. Additional, significant energy demand results from the electricity to be used primarily for reactor heating which is predicted to be between 8.6 and 9.2 kWh<sub>el</sub>/kg<sub>H<sub>2</sub></sub> [44].

#### 2.4. Carrier supply – CO<sub>2</sub> cycle

The return of the CO<sub>2</sub> to the place of the SNG production includes several steps being basically comparable to the components of a seaborne transportation of SNG. The storage density of the CO<sub>2</sub> captured during SNG reforming is substantially increased by liquefaction; therefore, the liquid form is preferable for efficient large-scale shipping. Various combinations of pressure and temperature between the triple point (5.18 bar, –56.6 °C) and the critical point (73.8 bar, 31.1 °C) are possible for CO<sub>2</sub> liquefaction [58]; i.e., CO<sub>2</sub> cannot be in a liquid state at pressures below 5.18 bar. Liquid CO<sub>2</sub> is already being transported by ship today. Current CO<sub>2</sub> carriers, primarily operated by the food and beverage industry, attain maximum transportation capacities of approx. 1,800 t CO<sub>2</sub> [59]. The tanks of these ships keep the CO<sub>2</sub> in a liquid state at relatively high pressures (16 to 21 bar) and temperatures between –22 and –30 °C [60,61]. However, liquid CO<sub>2</sub> shows its maximum storage density of 1,177.9 kg/m<sup>3</sup> at the triple point [62]. For carriers with much higher capacities, most likely needed for an economic CO<sub>2</sub> shipping in the context of industrial CO<sub>2</sub> cycles, storage conditions should be as close as possible to the triple point to allow maximum utilization of the available cargo capacity [63,64].

Taking into account the transportation conditions described above, only tanks that can both minimize heat entry and can be pressurized are suitable for shipping of liquid CO<sub>2</sub>. Such tanks, referred to as “semi-refrigerated” tanks, are known from shipping of liquefied petroleum gas (LPG) and ethylene; i.e., the development of large-scale CO<sub>2</sub> tankers may be based on the design of existing LPG and ethylene carriers [63]. The most ambitious concept for large-scale CO<sub>2</sub> carriers proposed so far envisages a transportation capacity of 91,000 m<sup>3</sup> (equals ~ 105,000 t<sub>CO<sub>2</sub></sub>) [65]. The fuel consumption of a vessel of such size would be in the range of 850 kWh<sub>LHV</sub>/km [61].

An additional electrical energy requirement for CO<sub>2</sub> recycling arises from liquefaction varying between 0.1 and 0.2 kWh/kg<sub>CO<sub>2</sub></sub> [65–67], whereby the lower bound appears feasible for newly build large-scale plants. The energy demand for loading and unloading the ship as well

as for CO<sub>2</sub> regasification is negligible in relation to shipping and liquefaction [64].

#### 2.5. Carrier supply – Supply of non-fossil CO<sub>2</sub>

Both, in the context of the SNG supply chain (methanation, liquefaction, transportation, regasification) and in the cycling of CO<sub>2</sub>, it is highly likely that small gas losses occur [64,68,69]. The respective carbon loss must be compensated – for reasons of climate protection and to maintain the supplied energy over time. To ensure a GHG-neutral and renewable energy supply, the loss compensation has to be based on non-fossil carbon. Therefore, two main options can be considered for a sustainable compensation of carbon losses:

- Feeding biogenic CO<sub>2</sub> into the CO<sub>2</sub> cycle
- Feeding atmospheric CO<sub>2</sub> into the CO<sub>2</sub> cycle

Biogenic CO<sub>2</sub> is released by various processes converting biomass into energy, a bio-energy carrier and/or a biomass-based raw material or product. Bioenergy processes with a substantial potential for CO<sub>2</sub> supply are biogas and bioethanol production as well as solid biofuel combustion. The effort of such a CO<sub>2</sub> capture are highly dependent on the respective source [70].

- The lowest effort arises if CO<sub>2</sub> from bioethanol plants is used as the gas produced during alcoholic fermentation consists almost exclusively of CO<sub>2</sub>. Accordingly, to make CO<sub>2</sub> from bioethanol production available for a CO<sub>2</sub> cycle in the sense of the concept discussed here consists solely in compression and transportation of the gas [71].
- Biogas contains a CO<sub>2</sub> concentration between 40 and 45 % [72]. The separation of this CO<sub>2</sub> from the biogas is common practice today when biogas is upgraded to biomethane. For this purpose, various processes are available (e.g., PSA, water scrubbing, membrane processes, cryogenic processes, amine scrubbing) [72,73]. Depending on the technology used, thermal and/or electrical energy is required [74].
- Also, biomass combustion processes, for example in heating plants burning wood fuels, can be used as a source for biogenic CO<sub>2</sub>. In conventional combustion with air, the CO<sub>2</sub> is diluted in the flue gas stream in concentrations between 8 and 15 vol-% [71]. Although CO<sub>2</sub> capture from combustion flue gases has not yet been applied for the supply of biogenic CO<sub>2</sub>, chemical (e.g., amine scrubbing) and physical removal processes have been proven on a large scale at power plants operated with fossil fuels [72]. Compared to biogas upgrading, where similar technical processes are used, a higher energy input has to be expected due to the lower CO<sub>2</sub> concentrations in the flue gas [70,71].

Direct air capture (DAC) is a process that enables the direct extraction of CO<sub>2</sub> from the ambient air. In this process, air is fed through an absorption unit that binds the CO<sub>2</sub>. As soon as the absorbent is saturated, the CO<sub>2</sub> is re-dissolved through heat supply within a regeneration process and is afterwards available as a pure stream [26]. Currently, different DAC processes are under development varying with regard to the sorbents used, the temperature level required for redissolution and the process management. Low-temperature processes typically use amine-based sorbents and require temperatures in the range of 100 °C [75]. High-temperature DAC employs an aqueous potassium hydroxide (KOH) or sodium hydroxide (NaOH) solution to absorb CO<sub>2</sub>; temperatures of approx. 900 °C are necessary to obtain the CO<sub>2</sub> that is converted into calcium carbonate (CaCO<sub>3</sub>) during the process [76]. DAC technologies are already available in small scale applications today. These units are characterized by high energy demands and substantial space requirements. To capture atmospheric CO<sub>2</sub>, between 1,500 and 2,000 kWh/t<sub>CO<sub>2</sub></sub> of thermal energy and between 250 and 450 kWh/t<sub>CO<sub>2</sub></sub> of electrical energy are needed [19,75,77]. For the supply of relevant CO<sub>2</sub>

quantities by means of DAC, a significant upscaling of the current plant size is required in addition to further technological development, e.g., to reduce energy demands.

### 3. Assessment methodology

This paper aims to assess SNG-based supply of elementary, “green” hydrogen. The methodological approach follows the structure outlined in Fig. 2. Firstly, the system to be assessed is defined along with its boundaries; the definition of the hydrogen supply chains is a key part of this system description (section 4.2) and is carried out in line with the selection of sound framework assumptions (section 4.1) as well as suitable publicly available technical data (section 4.3 to 4.7). The subsequent system modelling, which is based on the selected framework assumptions and data, is divided into two stages. Initially, each single component of the supply chains is modelled on the basis of scientific laws and technical relationships using a black-box approach. Subsequently, the components are linked to each other in order to model the entire supply chain. This modelling describes mass balances (input and output streams) and energy flows along the overall supply chain. Based on this system modelling, the SNG-based hydrogen supply chains can be evaluated with regard to technical, economic and environmental criteria. The following sections describe the mathematical approach for determining these evaluation criteria.

#### 3.1. Technical assessment

The energy efficiency being a main technical characteristic value describes the ratio between the energy provided by a process / process chain and the total energy expended for it. Correspondingly, the overall energy efficiency of the point-to-point hydrogen supply chain ( $\eta_{SC}$ ) is calculated by division of the energy content of the hydrogen supplied at the point-of-import ( $q_{H2\ at\ IP}$ ) by the total energy input along all supply chain components ( $q_{input}$ ) (equation (7)).  $q_{input}$  includes all forms of energy (e.g., electrical, thermal and chemical energy), irrespective of their exergy. The lower heating value (LHV) is the reference for the energy content of the hydrogen supplied and all chemical energy carriers used along the supply chain (e.g., shipping fuel).

$$\eta_{SC} = q_{H2\ at\ IP} / q_{input} \quad (7)$$

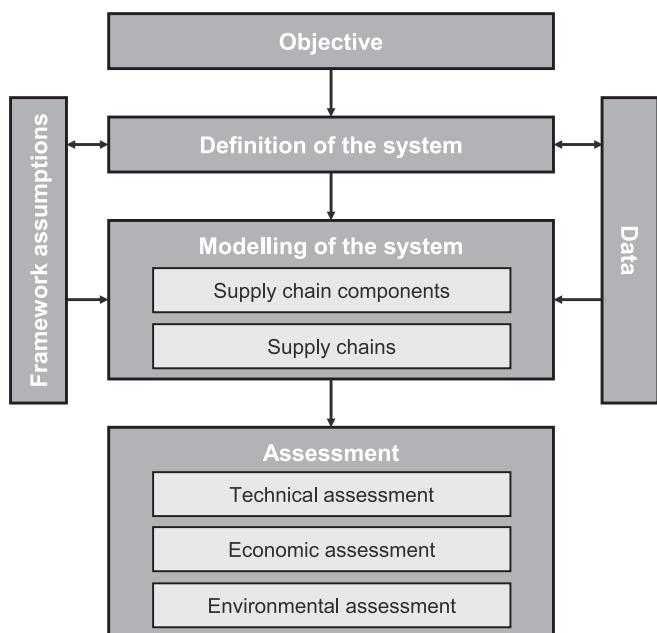


Fig. 2. General structure of the methodological approach.

The total energy input along the supply chain  $q_{input}$  is calculated as the sum of energy inputs for each component of the chain (equation (8)). These energy inputs of the specific supply chain components result from a black-box consideration and are based on literature data. Thermodynamic correlations within the individual components (e.g., during methanation) are therefore not considered in this work.  $q_{H2\ at\ EP}$  defines the energy input for a continuous supply of “green” hydrogen at the methanation plant resp. the point-of-export. The energy required for a continuous  $CO_2$  supply to the methanation plant is expressed as  $q_{CO2\ supply}$ . Additionally, the energy demands for methanation and SNG liquefaction ( $q_{conversion}$ ), for LSNG transportation via ship ( $q_{transport}$ ), LSNG storage ( $q_{storage}$ ) and the reconversion of LSNG into compressed, purified hydrogen ( $q_{reconversion}$ ) are considered.

$$q_{input} = q_{H2\ at\ EP} + q_{CO2\ supply} + q_{conversion} + q_{transport} + q_{storage} + q_{reconversion} \quad (8)$$

The total demand of hydrogen to be supplied to the methanation and thus the corresponding energy input for “green” hydrogen supply at the point-of-export  $q_{H2\ at\ EP}$  is largely influenced by losses that occur along the supply chain (e.g., boil-off losses during transportation, (re-)conversion losses). Therefore,  $q_{H2\ at\ EP}$  can be described as the sum of the energy that is needed for the supply of hydrogen effectively delivered at the point-of-import ( $q_{H2\ at\ EP, ideal}$ ) and the energy required to provide additional hydrogen needed to compensate the losses along the supply chain ( $q_{H2\ SC\ losses}$ ) (equation (9)).

$$q_{H2\ at\ EP} = q_{H2\ at\ EP, ideal} + q_{H2\ SC\ losses} \quad (9)$$

$q_{H2\ SC\ losses}$  includes, as summarized in equation (10), energy losses occurring during conversion from hydrogen to methane ( $q_{conversion\ loss}$ ), during the respective reconversion via methane reforming ( $q_{reconversion\ loss}$ ) as well as during SNG storage ( $q_{storage\ loss}$ ) and transportation ( $q_{transport\ loss}$ ). The latter two are mainly caused by boil-off. The conversion losses  $q_{conversion\ loss}$  describe, in a simplified form, the difference between the enthalpy of the educts (hydrogen and  $CO_2$ ) and the enthalpy of the product (SNG), and thus equal the heat generation during methanation. Additionally, energy losses caused by SNG leakage during its liquefaction are included in  $q_{conversion\ loss}$ . Correspondingly, the reconversion losses  $q_{reconversion\ loss}$  are composed of the difference between the enthalpy of the SNG input and the hydrogen output of methane reforming and the hydrogen losses during final compression.

$$q_{H2\ SC\ losses} = q_{conversion\ loss} + q_{transport\ loss} + q_{storage\ loss} + q_{reconversion\ loss} \quad (10)$$

Within the scope of the supply chains considered here, the total amount of  $CO_2$  to be supplied to the methanation plant can be divided into a part provided via a  $CO_2$  cycle and a part additionally supplied to compensate for the occurring carbon losses. Such carbon losses can result from incomplete  $CO_2$  capture at the reforming plant as well as from gas leakage during liquefaction, transportation and regasification of SNG and  $CO_2$ . Consequently, the total energy input for  $CO_2$  supply ( $q_{CO2\ supply}$ ) is calculated with equation (11).

$$q_{CO2\ supply} = q_{CO2\ supply, cycle} + q_{CO2\ supply, onsite} \quad (11)$$

A distinction is made between supply chains with and without a  $CO_2$  cycle (further details in section 4.2). In supply chains including a  $CO_2$  cycle,  $q_{CO2\ supply, onsite}$  corresponds to the energy required to compensate for the carbon losses described above by means of an onsite supply of non-fossil  $CO_2$ . In supply chains without a  $CO_2$  cycle,  $q_{CO2\ supply, onsite}$  equals the energy required to provide the total amount of  $CO_2$  needed for methanation via onsite supply. Correspondingly,  $q_{CO2\ supply, cycle}$  is only relevant for supply chains including  $CO_2$  cycling and describes the total energy demand associated with the return of the  $CO_2$  from the point of SNG reforming to the point of methanation. According to

equation (12),  $q_{\text{CO}_2 \text{ supply, cycle}}$  describes the energy demand for CO<sub>2</sub> liquefaction ( $q_{\text{CO}_2 \text{ liquefaction}}$ ), for CO<sub>2</sub> transportation ( $q_{\text{CO}_2 \text{ transport}}$ ) and CO<sub>2</sub> storage ( $q_{\text{CO}_2 \text{ storage}}$ ).

$$q_{\text{CO}_2 \text{ supply, cycle}} = q_{\text{CO}_2 \text{ liquefaction}} + q_{\text{CO}_2 \text{ transport}} + q_{\text{CO}_2 \text{ storage}} \quad (12)$$

### 3.2. Economic assessment

For the economic comparison of the different supply chains, the hydrogen supply cost at the point-of-import is used as the main comparative parameter. To do so, all economic assumptions and input parameters are first adjusted to the value of the euro (€) related to the year 2020 (€<sub>2020</sub>). The subsequent method for determining the hydrogen supply cost at the point-of-import ( $c_{\text{H}_2 \text{ at IP}}$ ) corresponds basically to the calculation of total energy input ( $q_{\text{input}}$ ). Thus,  $c_{\text{H}_2 \text{ at IP}}$  is the sum of the costs for all chain components, including cost for continuous hydrogen supply at the point-of-export ( $c_{\text{H}_2 \text{ at EP}}$ ), CO<sub>2</sub> supply ( $c_{\text{CO}_2 \text{ supply}}$ ), conversion from hydrogen to liquefied SNG ( $c_{\text{conversion}}$ ), transportation of SNG ( $c_{\text{transport}}$ ), storage of SNG ( $c_{\text{storage}}$ ) and final reconversion from SNG to compressed, purified hydrogen ( $c_{\text{reconversion}}$ ) (equation (13)).

$$c_{\text{H}_2 \text{ at IP}} = c_{\text{H}_2 \text{ at EP}} + c_{\text{CO}_2 \text{ supply}} + c_{\text{conversion}} + c_{\text{transport}} + c_{\text{storage}} + c_{\text{reconversion}} \quad (13)$$

$c_{\text{H}_2 \text{ at EP}}$  includes the cost incurred to supply hydrogen effectively delivered at the point-of-import and hydrogen needed to compensate hydrogen losses occurring along the supply chain. According to equation (14),  $c_{\text{CO}_2 \text{ supply}}$  includes the costs of the CO<sub>2</sub> cycle components and the costs of the onsite CO<sub>2</sub> supply required to compensate for carbon losses ( $c_{\text{CO}_2 \text{ supply, onsite}}$ ). With regard to the cost of the CO<sub>2</sub> cycle, liquefaction ( $c_{\text{CO}_2 \text{ liquefaction}}$ ), ship transportation ( $c_{\text{CO}_2 \text{ transport}}$ ) and intermediate storage ( $c_{\text{CO}_2 \text{ storage}}$ ) are taken into account.

$$c_{\text{CO}_2 \text{ supply}} = c_{\text{CO}_2 \text{ liquefaction}} + c_{\text{CO}_2 \text{ transport}} + c_{\text{CO}_2 \text{ storage}} + c_{\text{CO}_2 \text{ supply, onsite}} \quad (14)$$

The specific annual cost of the single components (i) being part of the SNG supply chain and the CO<sub>2</sub> cycle are calculated with equation (15). The total annual component cost is calculated by addition of the annual investment-related cost (ACAPEX<sub>i</sub>) and the annual operating cost (OPEX<sub>i</sub>). Through division by the amount of hydrogen provided at the point-of-import ( $\dot{m}_{\text{H}_2 \text{ at IP}}$ ) the specific cost of the single component ( $c_{\text{component, i}}$ ) is obtained. In OPEX<sub>i</sub>, energy related costs (e.g., fuel for shipping, electricity and heat for reforming or liquefaction) as well as costs for maintenance and repair are included.

$$c_{\text{component, i}} = \frac{\text{ACAPEX}_i + \text{OPEX}_i}{\dot{m}_{\text{H}_2 \text{ at IP}}} \quad (15)$$

By applying the annuity method, the annual investment-related cost ACAPEX<sub>i</sub> can be derived from the initial investment of the respective component (CAPEX<sub>i</sub>) taking into account the annuity factor (AF<sub>i</sub>) (equation (16)).

$$\text{ACAPEX}_i = \text{CAPEX}_i \text{ AF}_i \quad (16)$$

The annuity factor AF<sub>i</sub> is determined by equation (17). Here, the technology specific technical lifetime of the respective component (LT<sub>i</sub>) as depreciation period and the real weighted average cost of capital (WACC<sub>real</sub>) over that depreciation period are used.

$$\text{AF}_i = \frac{\text{WACC}_{\text{real}}(1 + \text{WACC}_{\text{real}})^{\text{LT}_i}}{(1 + \text{WACC}_{\text{real}})^{\text{LT}_i} - 1} \quad (17)$$

For the determination of WACC<sub>real</sub>, besides the nominal weighted average cost of capital (WACC<sub>nom</sub>), the projected future inflation rate (INFL) is also taken into account (equation (18)).

$$\text{WACC}_{\text{real}} = \frac{1 + \text{WACC}_{\text{nom}}}{1 + \text{INFL}} \quad (18)$$

### 3.3. Environmental assessment

The environmental assessment of the various hydrogen supply chains is limited to the overall GHG emissions. To account different types of GHG emitted and quantify their relative impact to climate change, the global warming potential of CO<sub>2</sub> for a time period over 100 years (GWP<sub>100</sub>) is used as a reference [78,79]. Therefore, GHG emissions caused by each supply chain component are taken into account for calculating the overall GHG emissions of the hydrogen supply at the point-of-import ( $e_{\text{H}_2 \text{ at IP}}$ ) (equation (19)).

$$e_{\text{H}_2 \text{ at IP}} = e_{\text{H}_2 \text{ at EP}} + e_{\text{CO}_2 \text{ supply}} + e_{\text{conversion}} + e_{\text{transport}} + e_{\text{storage}} + e_{\text{reconversion}} \quad (19)$$

According to equation (20), the specific GHG emissions of the chain components ( $e_{\text{component, i}}$ ) are composed of emissions caused by leakage of GHG ( $e_{\text{leakage, i}}$ ) and emissions through the use of energy ( $e_{\text{energy, i}}$ ). Emissions caused by leakage ( $e_{\text{leakage, i}}$ ) are only accounted for if they have an effective additional impact on global warming. In the concept analyzed here, only CO<sub>2</sub> of non-fossil origin is used to compensate for carbon losses. As a result, the CO<sub>2</sub> balance of the overall system is offset and there is no additional climate impact due to CO<sub>2</sub> leakage (e.g., through incomplete capture during methane reforming or slip during CO<sub>2</sub> liquefaction). This is contrasted by methane leakages (e.g., through methane slip during liquefaction or boil-off gas released to the atmosphere). While these are offset with regard to the carbon mass balance too, an additional climate impact must be considered due to the higher GHG potential of the released methane compared to the captured CO<sub>2</sub>.

$$e_{\text{component, i}} = e_{\text{leakage, i}} + e_{\text{energy, i}} \quad (20)$$

## 4. Framework, system definition and data

A fair technical, economic, and environmental assessment of SNG-based hydrogen supply chains is only possible with sound framework assumptions as well as realistic technology parameters set for the calculations. This section provides an overview of the framework defined, assumptions made and the technology parameters selected. The techno-economic data presented here and used for the system modelling is the result of comprehensive literature research. When more than two literature sources are cited for a single value, the median value of the cited sources was used.

### 4.1. Framework assumptions

A point-to-point supply is assumed for the evaluation of the different hydrogen supply chains. The point-of-import and –export are located in the immediate vicinity of the respective ports; i.e., international shipping is considered as the exclusive mode of transportation. The distance for shipping SNG from the point-of-export to the point-of-import and the return of CO<sub>2</sub> is 10,000 km; this is in line with the scale of frequently discussed [17,19,80–82] transportation routes for “green” energy (e.g., from Namibia, the Arabian Peninsula and the coast of Texas to ports in the North Sea like Rotterdam and Wilhelmshaven or also from the southern coast of Western Australia to Japan or Korea). For all examined supply chains, the supply of gaseous hydrogen with a purity ≥ 99.95 % and a pressure of 100 bar at the point-of-import is assumed.

For the calculation of the WACC<sub>real</sub> according to equation (18), the WACC<sub>nom</sub> is set to 8 % and the projected future inflation rate INFL to 2 % [83,84]. The reference year for all technology-specific parameters (technical, economic and environmental) is 2030. Operation with 8,000 h/a (full load) is presumed for all components downstream of the hydrogen supply at the point-of-export.

GHG emissions resulting from the construction and installation of supply chain components (e.g., PV plants, wind turbines, electrolyzers, methanation plants, ships) are not considered. This aligns with the methodology established by the European Union (EU) for assessing the GHG emissions associated with “green” hydrogen and other renewable fuels of non-biological origin (Commission Delegated Regulation (EU) 2023/1185) [85]. For the accounting of climate-impacting GHG emissions caused by gas leakage, a  $GWP_{100}$  of 25  $kg_{CO_2-eq}/kg$  is used for methane and 6  $kg_{CO_2-eq}/kg$  for hydrogen [86].

#### 4.2. Definition of hydrogen supply chains

The technology alternatives presented in sections 2.1 to 2.5 allow various possibilities for specific configurations of SNG-based hydrogen supply chains. For the definition of the point-to-point supply chains to be assessed according to the methodology introduced in section 3, a distinction is made between different technological concepts for the following supply chain components.

- **SNG reforming – ATR/eSMR**

While the thermal energy required for SNG reforming is provided internally when using the ATR (i.e., by using additional methane), the heat requirement for the eSMR is met externally by heating the reactor electrically.

- **Concept for CO<sub>2</sub> supply – With/without CO<sub>2</sub> cycle**

Supply chains with and without a CO<sub>2</sub> cycle are taken into account. If a CO<sub>2</sub> cycle is used, the CO<sub>2</sub> source available at the point-of-export is used exclusively to cover carbon losses. If no CO<sub>2</sub> is returned, the entire amount of required CO<sub>2</sub> is provided via DAC or from a biogenic source.

- **Available CO<sub>2</sub> source at the point-of-export – Biogenic source/DAC**

Supply pathways are considered that exploit a biogenic CO<sub>2</sub> source and pathways that extract CO<sub>2</sub> from ambient air by using the DAC technology.

Fig. 3 provides an overview of the SNG-based hydrogen supply chains and its main components. All possible combinations derived from the technologies discussed in section 2 result in a total of eight supply chains and are presented in Table 1. These supply chains will be assessed in this paper.

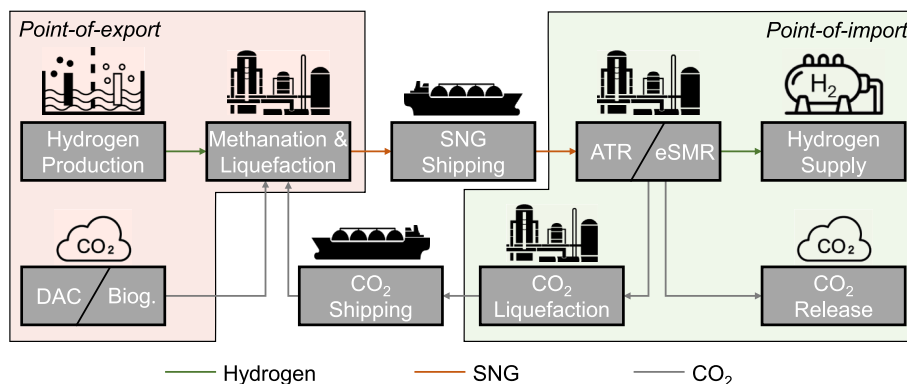
#### 4.3. Continuous hydrogen supply at the point-of-export

A continuous “green” hydrogen feed to the methanation unit is assumed. In addition to the water electrolysis, systems for a continuous “green” hydrogen supply usually also include seawater treatment, intermediate hydrogen storage and the necessary gas compression between the electrolyzer and the storage facility. Intermediate hydrogen storage is particularly relevant given the fluctuating supply of electricity from wind power and solar radiation. The detailed modeling of such a system for continuous hydrogen supply is not realized here. Instead, literature values are used to describe the technical, economic and environmental parameters of this continuous hydrogen supply.

For a continuous “green” hydrogen supply related to the year 2030 at the point-of-export, an energy input of 50  $kWh_{el}/kg_{H_2}$  is assumed corresponding to an efficiency of around 67 % in relation to the LHV [87,88]. The cost for such a continuous hydrogen supply is set at 3.5 €/kg<sub>H<sub>2</sub></sub> (based on [87]). Neglecting any construction and installation-related emissions (section 4.1), the system for “green” hydrogen supply causes GHG emissions exclusively through leakages. It is assumed that 0.2 % of the hydrogen leaks directly into the atmosphere during production [89]. Further climate-relevant hydrogen leakages occur during compression with losses of 0.5 % [88,90].

#### 4.4. Conversion

The production of SNG via catalytic methanation and its liquefaction are key processes for allowing the LSNG to be shipped; the respective parameters are listed in Table 2. Thus, the methanation plant achieves the maximum chemical conversion efficiency of 83.2 % (in Table 2



**Fig. 3.** SNG-based hydrogen supply chains with different technology alternatives (ATR = Autothermal reforming; Biog. = Biogenic carbon dioxide source; CO<sub>2</sub> = Carbon dioxide; DAC = Direct air capture; eSMR = Electrified steam methane reforming).

**Table 1**  
Configurations of the assessed hydrogen supply chains.

Supply Chain	Technology for SNG reforming	Concept for CO <sub>2</sub> supply	CO <sub>2</sub> source at the point-of-export
Internal Heat Supply – CO <sub>2</sub> Cycle – DAC	ATR	CO <sub>2</sub> cycle	Ambient air (via DAC)
Internal Heat Supply – CO <sub>2</sub> Cycle – Biogenic	ATR	CO <sub>2</sub> cycle	Biogenic CO <sub>2</sub> source
Internal Heat Supply – Onsite CO <sub>2</sub> Supply – DAC	ATR	CO <sub>2</sub> only from source at point-of-export	Ambient air (via DAC)
Internal Heat Supply – Onsite CO <sub>2</sub> Supply – Biogenic	ATR	CO <sub>2</sub> only from source at point-of-export	Biogenic CO <sub>2</sub> source
External Heat Supply – CO <sub>2</sub> Cycle – DAC	eSMR	CO <sub>2</sub> cycle	Ambient air (via DAC)
External Heat Supply – CO <sub>2</sub> Cycle – Biogenic	eSMR	CO <sub>2</sub> cycle	Biogenic CO <sub>2</sub> source
External Heat Supply – Onsite CO <sub>2</sub> Supply – DAC	eSMR	CO <sub>2</sub> only from source at point-of-export	Ambient air (via DAC)
External Heat Supply – Onsite CO <sub>2</sub> Supply – Biogenic	eSMR	CO <sub>2</sub> only from source at point-of-export	Biogenic CO <sub>2</sub> source

**Table 2**  
Techno-economic parameters for methanation and SNG liquefaction.

Parameter	Unit	Methanation	SNG liquefaction
Reference size	$t_{SNG}/h$	500	500
CAPEX	$M\epsilon_{2020}$	3,600 <sup>a</sup>	2,100 <sup>a</sup>
OPEX	$\%_{CAPEX}/a$	4	4
Lifetime	a	20	25
Electricity demand	$kWh_{el}/kg_{SNG}$	0.28 <sup>b</sup>	0.50 <sup>b</sup>
Heat generation	$kWh_{th}/kg_{SNG}$	2.86 <sup>b</sup>	0
Usable share of heat	$\%_{Heat\ generation}$	85	/
Hydrogen demand	$kg_{H2}/kg_{SNG}$	0.503 <sup>c</sup>	/
CO <sub>2</sub> demand	$kg_{CO2}/kg_{SNG}$	2.743 <sup>c</sup>	/
SNG loss	%	0	0.09

<sup>a</sup> Median value of literature data.

<sup>b</sup> Temperature level > 150 °C [27].

<sup>c</sup> Based on stoichiometrics [19,91].

depicted by the specific hydrogen and CO<sub>2</sub> demand) [19,91]. The energy bound in the hydrogen not converted into methane is released as heat and 85 % of that heat will be available for further use. The methane losses occurring during liquefaction are considered as leakage into the atmosphere (section 3.3).

To compensate for CO<sub>2</sub> losses along the CO<sub>2</sub> cycle or for supply chains without such a cycle, CO<sub>2</sub> supply at the methanation site via an onsite source can be realized by concepts utilizing biogenic CO<sub>2</sub> sources or with concepts applying the DAC technology (section 2.5). Table 3 shows the techno-economic parameters for CO<sub>2</sub> supply via DAC. A low-temperature DAC is considered allowing the integration of heat generated in methanation to cover resp. reduce the heat demand of the

**Table 3**  
Techno-economic parameter for CO<sub>2</sub> supply via low-temperature DAC.

Parameter	Unit	Low-temperature DAC
Reference size	$t_{CO2}/h$	150
CAPEX	$M\epsilon_{2020}$	830
OPEX	$\%_{CAPEX}/a$	4
Lifetime	a	25
Electricity demand	$kWh_{el}/kg_{CO2}$	0.25
Heat demand	$kWh_{th}/kg_{CO2}$	1.5 <sup>a</sup>

<sup>a</sup> Heat demand before integration of methanation heat; Temperature level ~ 100 °C [75].

**Table 4**  
Techno-economic parameter for LSN and LCO<sub>2</sub> carrier.

Parameter	Unit	LSNG carrier	LCO <sub>2</sub> carrier
Reference size	$m^3$	266,000	91,000
Payload	$t_{SNG}\ resp.\ t_{CO2}$	120,000 <sup>a</sup>	105,000 <sup>d</sup>
CAPEX	$M\epsilon_{2020}$	200	111
OPEX	$\%_{CAPEX}/a$	3	5
Lifetime	a	30	30
Average speed	km/h	30	30
Fuel demand	$kWh_{fuel}/km$	0 <sup>b</sup>	830
Boil-off rate	$\%_{Load}/d$	0.15	0.15
Engine slip	$\%_{BOG}$	0.18	/
Flash rate	$\%_{Load}$	0.13	0.13
Recovery rate flash	$\%_{Flash}$	95	95
Un/Loading time	h/transfer	96	45
Operating cost	$\epsilon_{2020}/h$	572	572
Harbor fees	$\epsilon_{2020}/transfer$	260,000	60,000
Utilization	h/a	8,000	8,000

<sup>a</sup> Storage density of 450 kg/m<sup>3</sup> assumed.

<sup>b</sup> Use of boil-off gas as fuel assumed, sufficient to meet energy demand of roundtrip.

<sup>c</sup> Value based on personal information from industry.

<sup>d</sup> Storage density of 1,150 kg/m<sup>3</sup> assumed.

process. If a CO<sub>2</sub> supply from biogenic sources is considered, unrestricted access to CO<sub>2</sub> from bioethanol and plants upgrading biogas to biomethane is assumed as both processes produce a high-quality CO<sub>2</sub> stream as a by-product (section 2.5). Thus, for the supply of biogenic CO<sub>2</sub> only an additional compression energy demand of 100 kWh/t<sub>CO2</sub> arises [97]. The specific cost for supplying biogenic CO<sub>2</sub> at the methanation site is estimated at 25  $\epsilon_{2020}/t_{CO2}$  [71].

Furthermore, methanation, SNG liquefaction and CO<sub>2</sub> supply can access a constant supply of renewable electricity in the case considered here. Accordingly, a high utilization of 8,000 h/a (full load) is assumed for these units. The specific cost of a baseload electricity supply largely relying on renewable energies in regions potentially exporting “green” energy is set at 0.07  $\epsilon_{2020}/kWh$  for the reference year 2030 [98]. The same specific cost is also assumed for a constant heat supply, which may be necessary to meet the thermal energy demand of the DAC unit not covered by methanation waste heat. The use of baseload electricity and heat is associated with GHG emissions of around 40 g<sub>CO2-eq</sub>/kWh.

#### 4.5. Ship transportation

The assumed parameters for SNG import by ship and the respective return of the CO<sub>2</sub> are shown in Table 4. The boil-off gas (BOG) generated during transportation is sufficient to power the LNG carrier, ensuring that no additional marine fuel supply is required. The LNG carrier is equipped with a contemporary two-stroke dual-fuel engine, where 0.18 % of the

BOG is released unburned as slip gas [69,99]. The flash rate describes the generation of additional BOG during the loading and unloading of tanks with cryogenic liquids. For capturing and recovering the BOG generated as flash during loading and unloading of the LNG carrier, a rate of 95 % is assumed [78]. At the port-of-export, this recovery involves the return to the SNG liquefaction plant; at the port-of-import, the BOG from the flash can be used directly in the reforming unit (ATR or eSMR).

At present, there is no empirical data available regarding the generation of BOG during the large-scale transportation of liquid CO<sub>2</sub> (LCO<sub>2</sub>) by ship. Thus, it is assumed that the boil-off rate for LCO<sub>2</sub> carriers as well as the flash and recovery rates do not differ significantly from that of LNG carriers [100]. However, an additional shipping fuel must be provided to power the LCO<sub>2</sub> carrier. A renewable fuel is considered, with costs of 0.11 €<sub>2020</sub>/kWh<sub>LHV</sub> and emissions of 85 gCO<sub>2</sub>/kWh<sub>LHV</sub>. There is no reliquefaction of the CO<sub>2</sub> occurring as boil-off during transportation.

#### 4.6. Storage

In order to balance the intermittent loading and unloading of ships with the near-continuous operation of the up- and downstream components, storage of LSNG and LCO<sub>2</sub> is necessary. Table 5 summarizes the techno-economic parameters considered for these storage systems. To determine the reference size of the storage systems a capacity equal to the ship's payload is assumed (storage factor of 1) [63]. The assumptions regarding the specific boil-off rates during LSNG and LCO<sub>2</sub> storage align

**Table 5**  
Techno-economic parameter for LSNG and LCO<sub>2</sub> storage tanks.

Parameter	Unit	LSNG storage tank		LCO <sub>2</sub> storage tank	
Reference size	m <sup>3</sup>	266,000 <sup>a</sup>		91,000 <sup>a</sup>	
CAPEX	M€ <sub>2020</sub>	540 <sup>b</sup>	[95,103]	250 <sup>b</sup>	[63,104]
OPEX	%CAPEX/a	2	[17]	5	[58,63]
Lifetime	a	30	[95]	30	based on [95]
Boil-Off rate	%/d	0.05	[68]	0.05	based on [68]
Recovery rate boil-off	%BOG	95	[68]	95	based on [68]
Average storage time	d	10 <sup>c</sup>	own assumption	4 <sup>c</sup>	own assumption

<sup>a</sup> According to carrier payload.

<sup>b</sup> Median value of literature data.

<sup>c</sup> Deviating storage times result from a different number of ships – since fewer ships are needed, longer, de/loading-free periods to be bridged by storage arise for SNG.

**Table 6**  
Techno-economic parameter for reconversion of SNG to hydrogen via ATR and eSMR.

Parameter	Unit	ATR		eSMR	
Reference size	t <sub>H2</sub> /h	200		200	
CAPEX	M€ <sub>2020</sub>	4,800 <sup>a</sup>	[39,45,50,51,92]	2,900 <sup>a</sup>	[39,44]
OPEX	%CAPEX/a	4	based on [39,51]	4	based on [39,44]
Lifetime	a	25	[45]	25	[39,44]
Methane demand	kg <sub>SNG</sub> /kg <sub>H2</sub>	3.1 <sup>a</sup>	[39,45,50,92]	2.5 <sup>a</sup>	[39,44]
Electricity demand	kWh <sub>el</sub> /kg <sub>H2</sub>	1.9 <sup>a</sup>	[39,45,50]	8.9	[39,44]
O <sub>2</sub> demand	kg <sub>O2</sub> /kg <sub>H2</sub>	3.1 <sup>b</sup>		/	
O <sub>2</sub> supply			based on [17,105–108]		
Electricity demand	kWh <sub>el</sub> /kg <sub>O2</sub>	0.29		/	
Cost	€ <sub>2020</sub> /kg <sub>O2</sub>	0.08		/	
CO <sub>2</sub> capture rate	%	95 <sup>a</sup>	[39,45,50,51,92]	95 <sup>c</sup>	
Outlet pressure	bar	30	based on [38,45]	30	based on [38,45]

<sup>a</sup> Median value of literature data.

<sup>b</sup> Based on stoichiometrics for the stated methane demand.

<sup>c</sup> assumed to be identical to ATR, since requirements for CO<sub>2</sub> capture seem to be similar.

with those applied to ship transportation. However, as the storage tanks under consideration are stationary systems in the immediate vicinity of the respective up- and downstream components, BOG recovery of 95 % is feasible [78]. The recovered BOG is either returned to the liquefaction or used directly in methanation (in the case of LCO<sub>2</sub> storage at point-of-export) resp. reforming (in the case of LSNG storage at point-of-import).

#### 4.7. Reconversion

Table 6 lists the techno-economic parameters of the reforming technologies considered for SNG reconversion into hydrogen and CO<sub>2</sub>. In order to ensure maximum CO<sub>2</sub> concentration in the product gas stream, ATR operation with pure O<sub>2</sub> is assumed. An air separation unit is considered to supply the required O<sub>2</sub>. The supplementary electrical energy demand of the ATR includes electricity required for the separation of the product gases (hydrogen and CO<sub>2</sub>) and other auxiliary units. In the case of the eSMR, the electricity required to externally heat the reactor is added.

After the SNG has been reformed, the produced hydrogen must be compressed to the targeted pressure level of 100 bar. In addition, the captured CO<sub>2</sub> must be liquefied to enable its shipping. Table 7 shows the techno-economic parameters of hydrogen compression and CO<sub>2</sub> liquefaction. Gas losses are not captured and leak directly into the atmosphere.

In line with the considerations for the export country, a constant

**Table 7**  
Techno-economic parameter for hydrogen compression and CO<sub>2</sub> liquefaction.

Parameter	Unit	Hydrogen compression	CO <sub>2</sub> liquefaction
Reference size	t <sub>H<sub>2</sub></sub> /h resp. t <sub>CO<sub>2</sub></sub> /h	200	1,500
CAPEX	M€ <sub>2020</sub>	192 <sup>b</sup>	430 <sup>b</sup> [58,63]
OPEX	% <sub>CAPEX/a</sub>	5 [88,90]	4 [63]
Lifetime	a	15 based on [88,90]	25 [17]
Electricity demand <sup>d</sup>	kWh <sub>el</sub> /kg <sub>H<sub>2</sub></sub> resp. kWh <sub>el</sub> /kg <sub>CO<sub>2</sub></sub>	0.6 based on [90,110,111]	0.12 <sup>b</sup> [17,58,63,65,97]
Hydrogen resp. CO <sub>2</sub> loss	%	0.5 [88,90]	0.1 <sup>c</sup> based on [68]

<sup>a</sup> For compression resp. liquefaction based on gas at 30 bar (outlet pressure of reforming).

<sup>b</sup> Median value of literature data.

<sup>c</sup> Assumed to be identical to SNG liquefaction.

electricity supply is also assumed for SNG reforming, final hydrogen compression and CO<sub>2</sub> liquefaction. Costs of 0.1 €<sub>2020</sub>/kWh are considered for the supply of baseload electricity in 2030 in regions which are expected to become import hubs for hydrogen (mainly Europe and East Asia) [80]. 85 g<sub>CO<sub>2</sub>-eq</sub>/kWh<sub>el</sub> is assumed for the emission intensity of this baseload electricity supply (based on [112,113]).

## 5. Results and discussion

In the following, the results of the hydrogen supply chain assessment are discussed. The results are based on a supply chain modeling according to the methodological approach introduced in section 3 and the data and assumptions set out in section 4. Fig. 4 shows the energy input, cost and GHG emissions for a SNG-based supply of elementary, gaseous hydrogen at 100 bar.

### 5.1. Energy input and hydrogen supply chain efficiency

The total energy input for the supply of one kWh<sub>LHV</sub> of hydrogen at the point-of-import ranges from 2.3 to 2.8 kWh, resulting in an overall energy efficiency between 36 and 44 %. The majority (80 to 90 %) of the energy demand is dedicated to hydrogen production at the point-of-export. While the choice of the CO<sub>2</sub> supply option has only a minor impact on the overall supply chain efficiency, the reforming technology exerts a more substantial influence. Thus, supply chains employing eSMR (i.e., meeting the heating demand of the reforming externally through additional electricity) are around five percentage points more efficient than those using ATR (i.e., meeting the heating demand of the reforming internally through consumption of additional methane). The reason is that the chemical reconversion efficiency (energy in methane to energy in hydrogen) of eSMR is clearly higher compared to ATR (Table 6). Accordingly, in supply chains with ATR considerably more SNG – and therefore more hydrogen – must be produced at the point-of-export to supply the same amount of hydrogen at the point-of-import. Due to this, in cases with ATR the energy input to compensate for the supply chain losses (*H<sub>2</sub> Production – Only Losses*) is about twice as high as in cases with eSMR, as shown in Fig. 4. The more efficient methane utilization in supply chains with eSMR goes hand in hand with an additional energy demand at the point-of-import for heat supply (*Reconversion – Only Heat Supply*). Overall, however, this form of external heat supply is more energy-efficient in the scenario under consideration than supply chains with ATR where additional methane is used to provide the heat internally.

With regard to the different CO<sub>2</sub> supply options considered, Fig. 4 shows that CO<sub>2</sub> supply using only a local biogenic source is most efficient and requires hardly any energy input relevant for the overall efficiency. This result is essentially based on the assumption that high-purity CO<sub>2</sub> is available, which only needs to be compressed for

methanation (section 4.4). The supply of the entire CO<sub>2</sub> demand via DAC leads to a drop in total efficiency of around three percentage points compared to the sole use of biogenic CO<sub>2</sub>. The energy demand of DAC-based CO<sub>2</sub> supply can be reduced by around two thirds through the implementation of a cycle. Taking into account the assumptions made regarding the CO<sub>2</sub> capture rate at the SNG reforming plant and other carbon losses (e.g., boil-off losses in the form of CO<sub>2</sub> and methane), around 90 % of the CO<sub>2</sub> required for methanation can be provided via the cycle. Given this high recycling rate, it is of secondary importance for the energy efficiency of supply chains with a CO<sub>2</sub> cycle whether the carbon losses are compensated by means of DAC or from a biogenic source. This is particularly true as the DAC's heat demand can be completely covered by methanation waste heat if the majority of the CO<sub>2</sub> is supplied via a cycle. This methanation heat is not sufficient to cover the many times higher heat demand of a CO<sub>2</sub> supply solely via DAC. Accordingly, the provision of additional heat (from external sources) is necessary in this case.

### 5.2. Hydrogen supply cost

The hydrogen supply costs in the considered pathways range between 6.6 and 8.6 €<sub>2020</sub>/kg<sub>H<sub>2</sub></sub>, as shown in Fig. 4. The composition of the costs largely corresponds to that of energy input. The hydrogen production has a predominant influence, accounting for 60 to 75 % of overall supply cost. Beyond hydrogen production, conversion (includes methanation and liquefaction), CO<sub>2</sub> supply, and reconversion contribute notably to the hydrogen supply cost, whereas SNG storage, SNG ship transportation, and hydrogen compression have only a marginal impact. Related to the costs of SNG transportation, a provision of shipping fuel via boil-off gas was assumed. Therefore, in Fig. 4 shipping fuel costs (as well as energy demand and GHG emissions) appear as part of *H<sub>2</sub> Production – Only Losses*.

The use of eSMR for reconversion leads to a reduction of total supply cost of about 1 €<sub>2020</sub>/kg<sub>H<sub>2</sub></sub> compared to reconversion by means of ATR. As with the energy input (section 5.1), the higher chemical conversion efficiency of the eSMR reduces the cost of producing additional hydrogen to compensate for supply chain losses (*H<sub>2</sub> Production – Only Losses*) by more than half. Further advantages for supply chains characterized by high chemical conversion efficiencies arise because fewer material flows (SNG and CO<sub>2</sub>) have to be liquefied, transported and temporarily stored in order to provide the same amount of hydrogen. For example, in supply chains with reconversion via eSMR, the components of the transportation chain can be dimensioned considerably smaller (e.g., capacity of the liquefaction plants, number of ships) – and thus the overall costs can be reduced – compared to chains with reconversion via ATR. Under the assumptions made, the outlined cost benefits of the supply chains with eSMR outweigh the additional costs of providing the reconversion heat via electricity (*Reconversion – Only Heat Supply*).

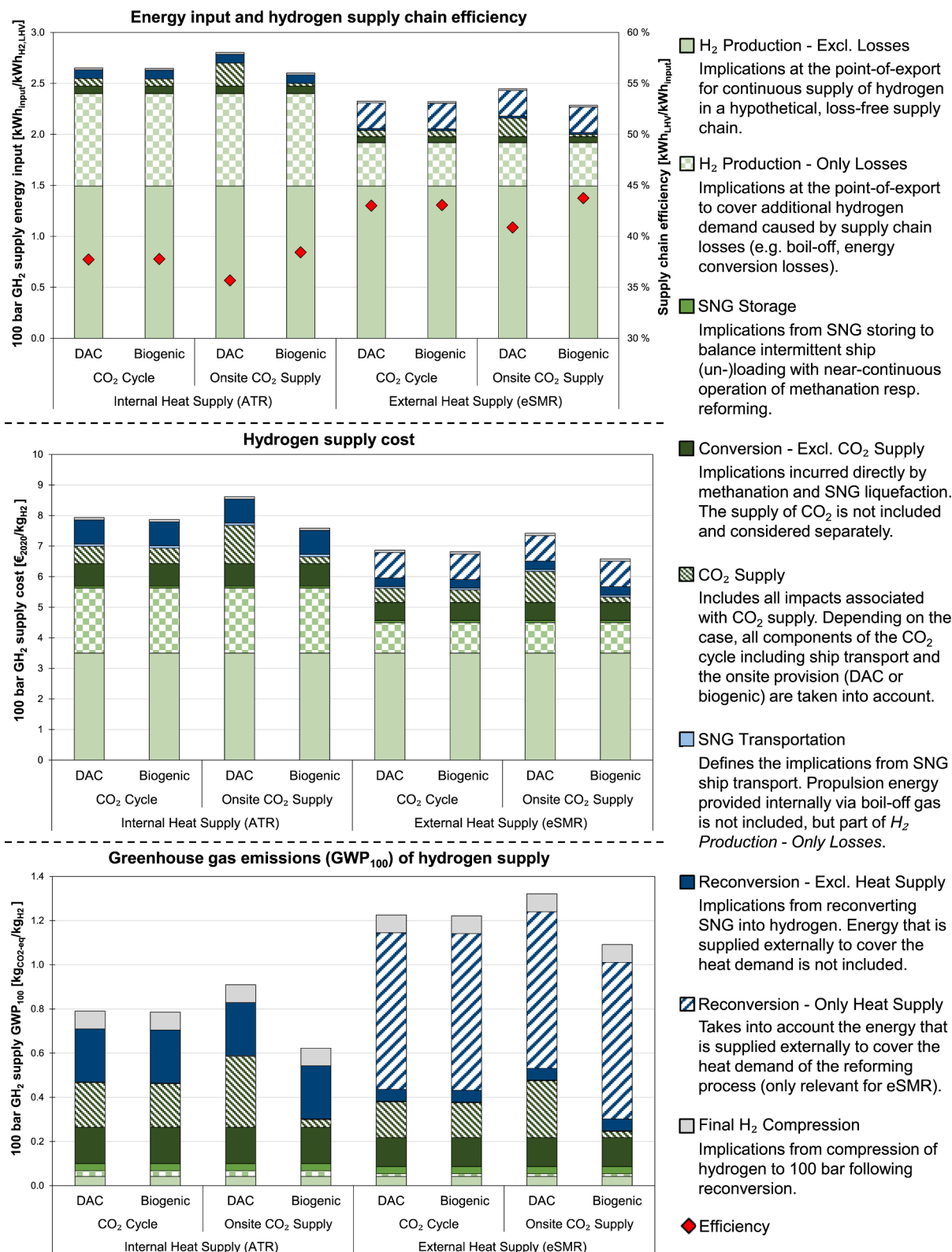


Fig. 4. Energy input and efficiency (top), cost (center) and GWP<sub>100</sub> (bottom) in 2030 for a SNG-based supply of gaseous hydrogen at 100 bar (ATR = Autothermal reforming; DAC = Direct air capture; eSMR = electrified steam methane reforming; GH<sub>2</sub> = Gaseous hydrogen; GWP<sub>100</sub> = Global warming potential for a time period over 100 years; LHV = Lower heating value, SNG = Synthetic natural gas).

Among the CO<sub>2</sub> supply options considered, supply chains utilizing a biogenic CO<sub>2</sub> source show the lowest costs, whereas supply exclusively via DAC leads to the highest costs. By implementing a CO<sub>2</sub> cycle, the CO<sub>2</sub> supply cost are more than halved compared to the sole use of DAC

systems; this leads to a reduction of the resulting hydrogen supply cost of approx. 0.7 €<sub>2020</sub>/kg<sub>H<sub>2</sub></sub> (if ATR is used for reconversion, for cases with eSMR slightly lower). Under the assumptions made, the overall costs of CO<sub>2</sub> cycling are around 60 €<sub>2020</sub>/t<sub>CO<sub>2</sub></sub>. At 40 €<sub>2020</sub>/t<sub>CO<sub>2</sub></sub>, ship

transportation is responsible for the majority of these costs, followed by the liquefaction with around 15 €<sub>2020</sub>/t<sub>CO2</sub>. The costs of the CO<sub>2</sub> cycle may be further reduced through the deployment of multi-gas carriers (transportation of LSNG and LCO<sub>2</sub> in the same ship) and the integration of SNG regasification cold into CO<sub>2</sub> liquefaction (these options were not assessed here). At 100 €<sub>2020</sub>/t<sub>CO2</sub>, the specific costs for CO<sub>2</sub> loss compensation using DAC are four times higher than the assumed costs for the supply from a biogenic source (section 4.1). Due to the high share of circulated CO<sub>2</sub> (~90 %), the availability of a biogenic CO<sub>2</sub> source has only little influence on the hydrogen supply cost if a CO<sub>2</sub> cycle is used.

For supply chains without such a cycle and with CO<sub>2</sub> supply exclusively onsite, the influence of biogenic CO<sub>2</sub> availability on the resulting hydrogen supply cost is significant. In that case, the DAC-based CO<sub>2</sub> supply incurs additional costs of 1 €<sub>2020</sub> (ATR case) resp. 0.85 €<sub>2020</sub> (eSMR case) per kg of hydrogen supplied at the point-of-import. As mentioned in section 5.1, in supply chains without a CO<sub>2</sub> cycle additional heat must be provided to cover the demand of the DAC. Thus, the specific cost for CO<sub>2</sub> supply by means of DAC is around 40 % higher when no CO<sub>2</sub> cycle is implemented (~140 €<sub>2020</sub>/kg<sub>CO2</sub>) compared to the previously discussed case (~100 €<sub>2020</sub>/kg<sub>CO2</sub>), where DAC is only used to compensate for carbon losses.

Against the background of the decisive importance of economic efficiency for a potential realization of SNG-based hydrogen supply, the sensitivity of the supply cost to a parameter variation is investigated.

The aim of this parameter variation is to map a broad variety of possible supply scenarios. Therefore, parameters are varied that are largely determined by the export and import regions and/or are significantly influenced by future techno-economic developments. To do so, the following parameters are assessed:

- Hydrogen production cost at the point-of-export
- Electricity cost at the point-of-import
- Transportation distance
- CO<sub>2</sub> supply cost at the point-of-export

For the variation the hydrogen production cost, electricity cost and transportation distance, no distinction between different CO<sub>2</sub> supply options is made; instead, a general onsite CO<sub>2</sub> supply cost of 25 €<sub>2020</sub>/t<sub>CO2</sub> is assumed. The influence of different point-of-export CO<sub>2</sub> cost on the resulting hydrogen supply cost is assessed separately by the variation of the corresponding parameter.

Fig. 5 shows the results of the parameter variation. It is apparent that a change in the point-of-export hydrogen production cost has a particularly strong influence on the resulting hydrogen supply cost. This impact of hydrogen production cost is slightly higher if the ATR is used to reconvert the SNG into hydrogen. The reason is that in the ATR the reforming heat is supplied internally and thus ultimately via additional hydrogen production at the point-of-export. If the eSMR is applied, the

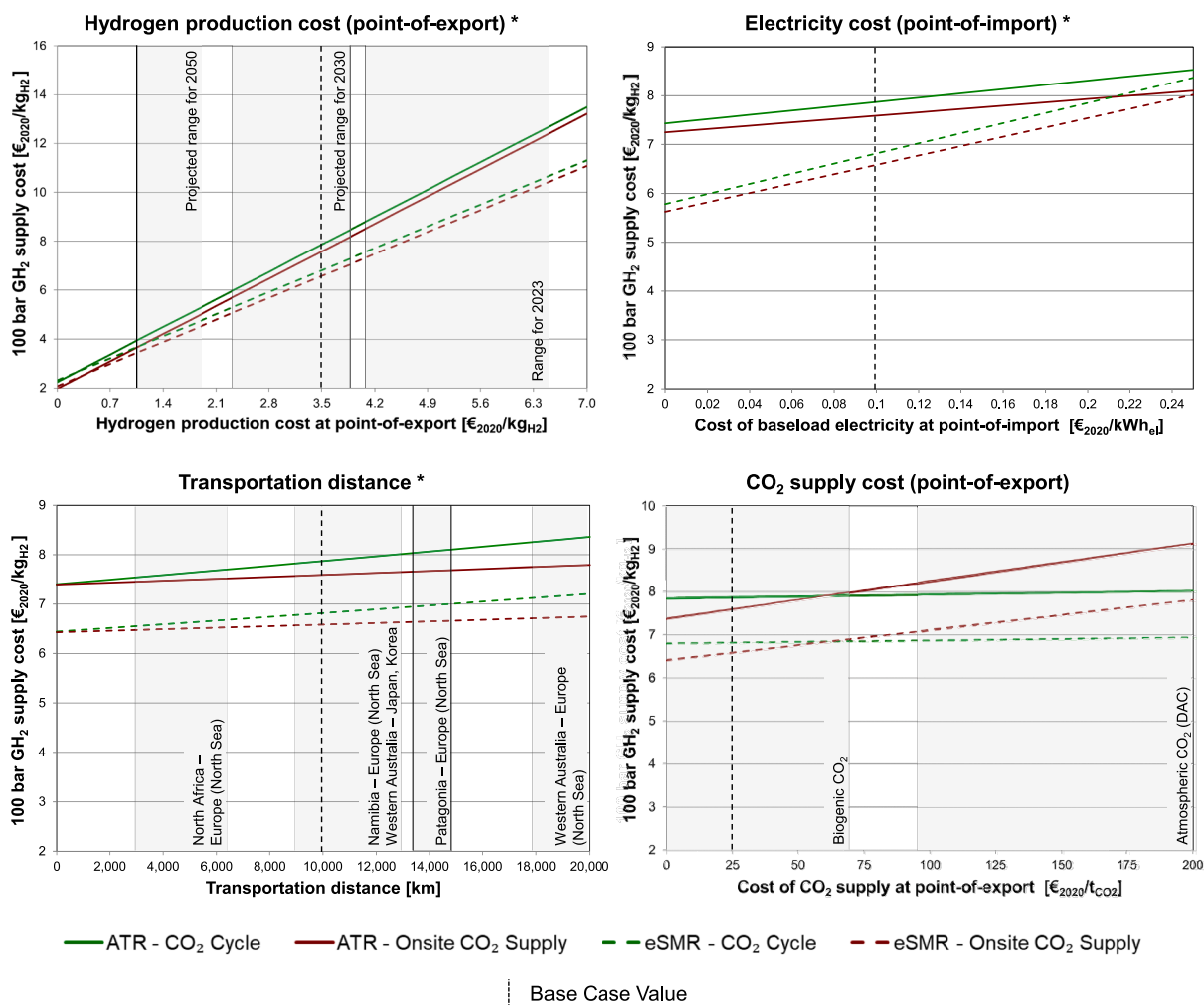


Fig. 5. Cost in 2030 for a SNG-based supply of gaseous hydrogen at 100 bar over variation of selected parameter

\* 25 €<sub>2020</sub>/t<sub>CO2</sub> assumed for CO<sub>2</sub> supply at point-of-export

(ATR = Autothermal reforming; DAC = Direct air capture; el = Electricity; eSMR = electrified steam methane reforming; GH<sub>2</sub> = Gaseous hydrogen; LHV = Lower heating value).

reforming heat is supplied by external electricity, resulting in a lower overall hydrogen demand and thus a lower dependence on its production cost. Furthermore, the variation in hydrogen production cost shows that the costs caused by transportation and the associated conditioning are around  $2 \text{ €}_{2020}/\text{kg}_{\text{H}_2}$ . Since the onsite  $\text{CO}_2$  supply cost ( $25 \text{ €}_{2020}/\text{t}_{\text{CO}_2}$ ) assumed in the scenario presented in Fig. 5 are lower than the specific  $\text{CO}_2$  recycling cost ( $\sim 60 \text{ €}_{2020}/\text{t}_{\text{CO}_2}$ ), the pathways with a  $\text{CO}_2$  cycle have slightly higher transportation-related costs than those with onsite supply. With regard to the influence of the electricity cost at the point-of-import, the results paint a contrary picture. Thus, due to their higher energy demand at the point-of-import the hydrogen supply cost of eSMR pathways show a higher dependency on the electricity cost than pathways with ATR. Moreover, supply pathways with a  $\text{CO}_2$  cycle are marginally more sensitive to the point-of-import electricity cost due to the additional electricity consumption of  $\text{CO}_2$  liquefaction. In summary, the choice between ATR and eSMR is mainly influenced by the difference in energy cost in the export and import region. Basically, the eSMR has advantages due to its higher overall efficiency. However, export scenarios in which hydrogen production in the export region is very cheap and the electricity supply in the import region is very expensive are favorable for the ATR, as energy is almost exclusively consumed at the point-of-export here. Here, the more efficient utilization of the produced hydrogen by eSMR is less impactful on the overall hydrogen supply cost. For example, in a hypothetical scenario in which hydrogen could be produced at the point-of-export for  $1.5 \text{ €}_{2020}/\text{kg}_{\text{H}_2}$ , the cost parity between supply pathways with eSMR and ATR would be given at point-of-import electricity cost of around  $0.17 \text{ €}_{2020}/\text{kWh}_{\text{el}}$ .

The lower charts of Fig. 5 illustrate the influence of the transportation distance and the point-of-export  $\text{CO}_2$  supply cost on the overall hydrogen supply cost. While the two parameters discussed in the previous paragraph mainly determine the cost difference between pathways with different SNG reforming technologies, the transportation distance and the  $\text{CO}_2$  cost primarily affect whether the use of a  $\text{CO}_2$  cycle is beneficial. In terms of transportation distance, the influence on the hydrogen supply cost is generally rather low. For example, doubling the distance of 10,000 km assumed in the base case only results in an increase in supply cost of less than  $0.5 \text{ €}_{2020}/\text{kg}_{\text{H}_2}$  across all considered pathways. Supply pathways that do not involve a  $\text{CO}_2$  cycle have a particularly low dependency on transportation distance. If a  $\text{CO}_2$  cycle is implemented, the  $\text{CO}_2$  transportation adds additional distance-related cost parameters (number of  $\text{CO}_2$  carrier required, fuel consumption), resulting in a slightly higher sensitivity. The point-of-export  $\text{CO}_2$  supply cost have hardly any influence on the hydrogen supply cost of pathways with a  $\text{CO}_2$  cycle under the given assumptions (the  $\text{CO}_2$  recycle rate calculated to around 90 % is particularly important here). However, if  $\text{CO}_2$  recycling is not used, high point-of-export  $\text{CO}_2$  cost result in high hydrogen supply cost. For a  $\text{CO}_2$  cost of  $200 \text{ €}_{2020}/\text{t}_{\text{CO}_2}$ , for example, the savings that can be realized through the  $\text{CO}_2$  cycle are around  $1 \text{ €}_{2020}/\text{kg}_{\text{H}_2}$ . It can be concluded that the cost advantages or disadvantages of implementing a  $\text{CO}_2$  cycle are always decisively influenced by the transportation distance and the  $\text{CO}_2$  cost at the point-of-export. For example, the cost parity between supply pathways with and without a  $\text{CO}_2$  cycle for a transportation distance of 3,000 km is around  $40 \text{ €}_{2020}/\text{t}_{\text{CO}_2}$ . If the distance were 20,000 km, as would be the case for exporting SNG from Australia to Europe, the  $\text{CO}_2$  costs at the point-of-export would have to exceed  $120 \text{ €}_{2020}/\text{t}_{\text{CO}_2}$  to make a  $\text{CO}_2$  cycle economically advantageous.

### 5.3. Greenhouse gas emissions of hydrogen supply

As Fig. 4 illustrates, the GHG emissions across the evaluated supply chains range from 0.6 to  $1.3 \text{ kg}_{\text{CO}_2\text{-eq}}/\text{kg}_{\text{H}_2}$ . In the considered framework, GHG emissions from hydrogen supply predominantly stem from baseload energy consumption during conversion from hydrogen to methane,  $\text{CO}_2$  provision, and reconversion processes. Leakage-related emissions, accounting for 15 to 30 % of overall emissions, are mainly caused by hydrogen production, initial and final hydrogen compression

as well as methane liquefaction. However, the sensitivity to the selected technology data is high. If, for example, a leakage rate of 1.5 % (corresponds to the range used by [66,96]) were assumed for methane liquefaction instead of 0.1 % ([68], see Table 2), the share of leakage-related emissions would rise to above 50 % while total emissions would increase to a range between 1.6 and  $2.1 \text{ kg}_{\text{CO}_2\text{-eq}}/\text{kg}_{\text{H}_2}$ . Furthermore, it is also conceivable that, for example due to technical reasons, the boil-off gas occurring on the LSNG tanker is not burned almost completely as assumed, but escapes into the atmosphere unused in parts. In this case, SNG transportation would become an additional relevant source of leakage-related emissions.

Notably, the level of GHG emissions is primarily influenced by the choice of reforming technologies, with supply chains using ATR demonstrating superior environmental performance with lower total GHG emissions compared to those using eSMR, despite the latter's higher energy efficiency. Thus, using ATR can reduce total emissions in the supply chains studied by 30–45 %. The higher overall energy efficiency of supply chains with eSMR has no decisive influence on the overall GHG emissions. Based on the assumption that the GHG emissions from hydrogen supply at the point-of-export are caused exclusively by leakage and are therefore comparatively low (section 4.3), the increased hydrogen losses from reconversion via ATR only lead to slight additional emissions. This is contrasted by the additional demand of baseload electricity by eSMR (*Reconversion – Only Heat Supply*), which, as assumed here, has potentially high specific emissions (section 4.7). If hydrogen supply at the point-of-export were to be accounted with higher specific GHG emissions, for example by taking into account the upstream emissions of renewable electricity generation, the different overall efficiency of supply chains with ATR and eSMR could have a noteworthy impact on the total GHG emissions. Thus, as already shown with regard to the costs, the emissions assessment is also largely dependent of the difference between emissions of energy supply in the import and export region. Low-emission hydrogen supply in export regions favors ATR-based supply chains, whereas eSMR benefits from cleaner electricity in import regions.

In the scenario under consideration,  $\text{CO}_2$  cycling can save emissions compared to onsite  $\text{CO}_2$  supply by means of DAC, as shown in Fig. 4. This is true, although the energy used for the  $\text{CO}_2$  cycle (mainly electricity for  $\text{CO}_2$  liquefaction at the point-of-import and renewable shipping fuel) has, based on the assumptions made, significantly higher specific emissions than the baseload electricity and heat used for DAC. However, this systemic advantage is outweighed by the many times higher total energy demand of the DAC compared to the  $\text{CO}_2$  cycle (section 5.1).

## 6. Conclusion

The goal of this paper is to assess elementary hydrogen supply by using SNG as a carrier for “green” hydrogen. As part of this assessment, novel approaches for the provision of the required  $\text{CO}_2$  as well as the reforming of methane to hydrogen are being investigated. To achieve this, this work evaluates the energy efficiency, cost, and GHG emissions of various supply concepts for the year 2030. The analysis of the technology components part of such supply chains illustrates that, from a technical point of view, hydrogen supply using SNG as a carrier can be a valid alternative to other ship-based hydrogen supply chains based on, e.g., ammonia, methanol or LOHC. Existing infrastructures and established technologies can be used for certain components of SNG-based hydrogen supply chains. However, there are significant technical challenges to overcome for a full-scale implementation. In particular, the catalytic methanation of hydrogen and  $\text{CO}_2$  as well as methane reforming for the supply of pure hydrogen with a simultaneous capture of  $\text{CO}_2$  have not yet been realized on a large scale. Also, for the supply of large quantities of non-fossil  $\text{CO}_2$ , technological innovations and/or a substantial scale-up of the current status quo may be necessary, especially if biogenic sources are not or not in the required quantity available at the point-of-export.

For a hydrogen supply based on SNG as a carrier, various supply pathways are defined, which differ in terms of the CO<sub>2</sub> supply option and the technical approach for the SNG reforming. Overall, the following conclusions can be drawn from the comparison of these supply pathways.

- The energy efficiency of SNG-based hydrogen supply, summing up to 36 to 44 % in the investigated pathways, is largely determined by hydrogen production at the point-of-export. The amount of hydrogen that has to be produced at the point-of-export, in turn, is clearly influenced by the supply chain losses to be compensated. Accordingly, the realization of a highly efficient energy conversion (electrolysis, methanation, SNG reforming) and low-loss energy transportation (e.g., boil-off) are essential for a SNG-based hydrogen supply with a high overall energy efficiency.
- For the evaluated concepts, the hydrogen supply cost varies between 6.6 and 8.6 €<sub>2020</sub>/kg<sub>H2</sub> related to the year 2030. This cost of SNG-based hydrogen supply is significantly affected by hydrogen production at the point-of-export and thus by the supply chain losses.
- Under the framework assumptions considered here, the climate-impacting GHG emissions from SNG-based hydrogen supply result largely from baseload energy use along the supply chain. Accordingly, the specific emissions of the electricity and heat used for methanation, SNG liquefaction, CO<sub>2</sub> supply and SNG reforming are of major importance for the overall emissions of such a hydrogen supply, which are between 0.6 and 1.3 kg<sub>CO2-eq</sub>/kg<sub>H2</sub>. However, this order of magnitude is only valid assuming that barely any gas leakages occur along the SNG transportation chain. Thus, the release of just 1.5 % of the produced SNG into the atmosphere (for example in the course of liquefaction or as boil-off during transportation) would lead to a substantial increase in GHG emissions.
- As part of the supply pathways comparison, ATR and eSMR are being investigated – two processes that enable high CO<sub>2</sub> capture rates during reconversion of SNG into hydrogen. This assessment shows that the eSMR is generally advantageous and allows a reduction of hydrogen supply cost by 10 to 20 % due to the more efficient utilization of the SNG; i.e., if the eSMR is used, less SNG is required to provide the same amount of hydrogen. Accordingly, less hydrogen needs to be produced at the point-of-export and, consequently, all downstream components of the supply chain can be dimensioned smaller. However, scenarios are conceivable in which systemic framework conditions for hydrogen imports arise that favor SNG reconversion via ATR. Thus, particularly high differences between import and export country energy cost (cheap hydrogen in export country and expensive electricity in import country) work in favor of ATR, as here almost all energy is required in the exporting country, whereas in the case of eSMR, a relevant share of total energy consumption occurs in the importing country.
- In order to enable a hydrogen supply that is climate-neutral, the CO<sub>2</sub> which is required for the conversion of hydrogen into SNG must be of non-fossil origin. One possible approach is to use the CO<sub>2</sub> within a closed cycle. A CO<sub>2</sub> supply by means of such a closed cycle potentially offers comparatively easy scalability, especially in comparison to the sole use of limited biogenic CO<sub>2</sub> sources. Whether the overall hydrogen supply cost can be reduced by using such a CO<sub>2</sub> cycle depends on various factors. The availability of a low-cost biogenic CO<sub>2</sub> source at the point-of-export and the transportation distance are of particular importance. From today's perspective, most likely such a CO<sub>2</sub> cycle leads to a cost reduction, if no biogenic CO<sub>2</sub> source is available at the point-of-export and the onsite supply would have to be realized via the DAC process. For a transportation distance of 10,000 km, for example, the specific cost of CO<sub>2</sub> cycling is around 60 €<sub>2020</sub>/kg<sub>CO2</sub> and thus significantly below the cost to be expected for CO<sub>2</sub> supply with DAC.

## Funding

This research did not receive any specific grant from funding agencies in the public, commercial, or not-for-profit sectors.

## CRedit authorship contribution statement

**Fabian Carels:** Writing – original draft, Visualization, Validation, Methodology, Investigation, Data curation, Conceptualization. **Lucas Sens:** Writing – review & editing, Supervision, Methodology, Data curation. **Martin Kaltschmitt:** Writing – review & editing, Supervision, Conceptualization.

## 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.

## Data availability

Data will be made available on request.

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