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Techno-economic and environmental assessment of renewable hydrogen import routes from overseas in 2030

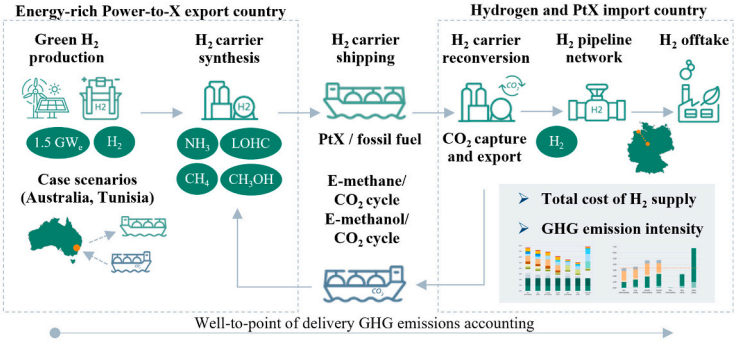
Florian Scheffler [a,*], Christoph Imdahl [a], Sabrina Zellmer [a,b], Christoph Herrmann [a,c]

[a] Fraunhofer Institute for Surface Engineering and Thin Films IST, Riedenkamp 2, Braunschweig 38108, Germany
[b] Technische Universität Braunschweig, Institute for Particle Technology, Volkmaroder Straße 4, Braunschweig 38104, Germany
[c] Technische Universität Braunschweig, Institute of Machine Tools and Production Technology, Langer Kamp 19b, Braunschweig 38106, Germany

HIGHLIGHTS

- CO_2/e-methane and CO_2/e-methanol based on a CO_2 transport cycle with CCUS.
- GHG emissions intensity (well-to-wake) for fossil and green marine fuels.
- Ammonia achieves the lowest hydrogen supply cost after reconversion of the carrier.
- Ammonia and methane comply with the GHG emissions treshold for long-distance routes.

GRAPHICAL ABSTRACT



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ABSTRACT

Converting renewable electricity via water electrolysis into green hydrogen and hydrogen-based products will shape a global trade in power-to-x (PtX) products. The European Union's renewable hydrogen import target of 10 million tonnes by 2030 reflects the urgent need for PtX imports by sea to early high-demand countries like Germany. This study evaluates the cost efficiency and greenhouse gas (GHG) emissions of four hydrogen carrier ship import options considering a reconversion to H_2 at the import terminal for a final delivery to offtakers via a H_2 pipeline network in 2030. This includes ammonia, a liquid organic hydrogen carrier (LOHC) system based on benzyltoluene (BT) and a novel CO_2/e-methane and CO_2/e-methanol cycle, where CO_2 is captured at the reconversion plant and then shipped back to the PtX production site in a nearly closed carbon loop. The GHG emission accounting includes well-to-wake emissions of the marine fuels and direct emissions of the carbon capture plant. Two GW-scale case studies reveal the impact of a short and long-distance route from Tunisia and Australia to Germany, whereas the specific PtX carriers are either fuelled by its PtX cargo as a renewable marine fuel or by conventional heavy fuel oil (HFO). Ammonia outperforms the other PtX routes, as the total hydrogen supply cost range between 5.07 and 7.69 for Australia (low: NH_3 HFO, high: LOHC HFO) and 4.78–6.21 € per kg H_2 for Tunisia (low: NH_3 HFO, high: CH_4 HFO), respectively. The ammonia routes achieve thereby GHG intensities of 31 % and 86 % below the EU threshold of 3.4 kg CO_2(e) per kg H_2 for renewable hydrogen. LOHC though, unless switching to low-emission fuels, and the CO_2/e-methanol cycle exceed the GHG threshold at

* Corresponding author.
E-mail address: florian.scheffler@ist.fraunhofer.de (F. Scheffler).

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shipping distances of 12,300 and 16,600 km. The hydrogen supply efficiencies vary between 57.9 and 78.8 $\%_{LHV}$ (low: CH_4 PtX-fuelled, high: NH_3 HFO) with a PtX marine fuel consumption of up to 15 $\%_{LHV}$ for the Australian methanol route, whereas high uncertainties remain for the ammonia and methanol reconversion plant efficiencies. The CO_2 cyle enables a cost-efficient CO_2 supply easing the near-term shortage of climate-neutral CO_2 sources at the cost of high GHG emissions for long-distance routes.

Nomenclature

ASU	Air separation unit
ATR	Autothermal reforming
AWE	Alkaline water electrolyser
BECCS	Bioenergy with carbon capture and storage
BOG	Boil-off gas
BT	Benzyltoluene
CAPEX	Capital expenditure
CCUS	Carbon capture, utilization and storage
CSIRO	Commonwealth Scientific and Industrial Research Organisation
DAC	Direct air capture
DBT	Dibenzyltoluene
DME	Dimethyl ether
DRI	Direct-reduced iron
e-NG	Electric natural gas (synthetic methane)
EHB	European Hydrogen Backbone
FLH	Full load hours (per year)
FT	Fischer-Tropsch
GHG	Greenhouse gas (GHG) (emissions)
GWP100	100-year time-horizon global warming potential
HB	Haber–Bosch
HFO	Heavy fuel oil
HHV	Higher heating value
IEA	International Energy Agency
IMO	International Maritime Organisation
IRENA	International Renewable Energy Agency
LCOE	Levelized cost of electricity
LCOH	Levelized cost of hydrogen
LHV	Lower heating value
LNG	Liquified natural gas
LOHC	Liquid organic hydrogen carrier
LPG	Liquefied petroleum gas
MSR	Methanol steam reforming
OPEX	Operational expenditure
PEMWE	Proton-exchange membrane (PEM) water electrolyser
PtX	Power-to-X
PSA	Pressure swing adsorption
PV	Photovoltaic
RE	Renewable electricity
RED	Renewable Energy Directive
RFNBO	Renewable fuels of non-biological origin
SMR	Steam methane reforming
SNG	Synthetic natural gas
SWRO	Seawater reverse osmosis
VRE	Variable renewable energies

1. Introduction

Integrating hydrogen as a versatile clean molecule is a fundamental pillar of the energy transition to defossilise the global energy system. [1] The use of hydrogen and hydrogen-based products is thereby prioritized for the reduction of greenhouse gas emissions (GHG) in hard-to-abate sectors, where direct electrification is unfeasible and energy-rich molecules are required, such as chemicals, steel, international shipping and aviation or seasonal energy storage. [2,3] A rapid production ramp-up of clean hydrogen is therefore needed from less than 1 million tons (Mt) of the current global production of 95 Mt. in 2022, towards 154 and 614 Mt. in 2030 and 2050 as demand forecasts in IRENA's 1.5 °C Scenario. [2,4] The majority of this new production is predicted to come from water electrolysis plants consuming exclusively renewable electricity (RE) to split water into hydrogen and oxygen referred to as 'green hydrogen'. The global installed capacity of electrolysis plants needs to grow accordingly from around 700 MW in 2022 to 350 GW in 2030 or 4–5 TW in 2050, as the number of announced projects rises steadily with a potential global capacity of up to 420 GW in 2030. [2–4] Green hydrogen can then be converted into multiple fuels and feedstocks, while all molecules derived from electricity are comprised as power-to-x (PtX) products. Additionally, low-carbon hydrogen will be sourced from processing fossil fuels, mostly natural gas (NG), with carbon capture, utilization and storage (CCUS). An accelerated global expansion of RE generation in countries with an abundant potential of variable renewable energies (VRE) from solar photovoltaic (PV) and wind power, is thereby the key enabler of the emerging hydrogen market based on a cost-effective green hydrogen production as electricity cost continue to decline and gigawatts of electrolysis plants are deployed. [5] A global market for green hydrogen and PtX products is predicted as multiple countries comprising today's leading fossil-fuel exporters as well as new energy exporters utilize their technical low-cost production potentials for renewable energy exports to high-demand countries. [6]

Hydrogen's pivotal role in the global energy transition is reflected by multiple national hydrogen strategies that were published over the last years with plans to export or import hydrogen dependent on their domestic production potentials from renewable energy sources and their demand for energy imports. [6] The European Union's hydrogen strategy sets the target of an installed electrolysis capacity of 40 GW in the EU by 2030 and has been complemented by the REPowerEU plan in 2022 by adding 10 million tons of renewable hydrogen imports to the domestic production target of 10 million tons by 2030. [7,8] Germany's hydrogen strategy was updated in 2023 with reinforced policies and targets, such as increasing the electrolysis target to 10 GW, deploying a dedicated hydrogen transmission network with more than 1800 km of pipelines by 2028, and accelerating the import of hydrogen and hydrogen derivatives by ship and pipeline in particular since imports already need to meet up to 70 % of the hydrogen demand in 2030. [9] In this context Germany revealed its import strategy for hydrogen and hydrogen derivatives in 2024 particularly addressing the dimensions and measures for the ramp-up of diversified hydrogen import routes by pipeline and ship transport, while first auctions for PtX imports are already placed in the H2Global funding and trading instrument and first contracts were awarded for large-scale green ammonia imports from Egypt in 2027. [9–11]

Current infrastructure projects include the deployment of the hydrogen pipeline core network integrated into the pan-European hydrogen pipeline network with its pipeline-based import corridors and the implementation of import terminals for hydrogen derivatives and PtX products. [10] The five import corridors of the European Hydrogen Backbone are an essential part of Europe's hydrogen supply

with cross-border pipelines planned for example from North Africa via Tunisia to Italy and Germany or between Portugal, Spain, France and Germany. [12,13] Ship-based hydrogen imports are moreover planned in Belgium, the Netherlands or Germany with for instance import terminals and reconversion plants for ammonia and synthetic methane at the Port of Wilhelmshaven, partly utilizing new liquefied natural gas (LNG) terminals, to deliver hydrogen via pipeline to early industrial offtakers, e.g. in the steel or chemical sector. [10,14–16]

Establishing global hydrogen supply chains by ship at a multi-megawatt-scale until 2030 is the main focus of this study and requires the conversion of hydrogen into a liquid hydrogen carrier, such as ammonia (NH_3), methanol (CH_3OH), methane (CH_4), Fischer-Tropsch (FT) derived jet fuels or a liquid organic hydrogen carrier (LOHC). Liquid hydrogen carrier can utilize existing infrastructure from fossil fuel markets with adequate capacities for storage and ship transportation. Moreover, these PtX products offer an increased volumetric energy density compared to compressed or liquified hydrogen. [17] Shipping liquid hydrogen (LH_2) still phases various challenges as current liquefaction and tank storage capacities need to be scaled up multiple times to achieve nearly equal energy equivalents as for the transport of ammonia, liquefied natural gas (LNG) or methanol. [17,18] Hydrogen has the second lowest boiling point of all elements at −253 °C (1 bar) leading to an energy-intensive liquefaction process and material challenges for storage as the entire value chain is less mature and developed than handling for instance LNG under cryogenic conditions at −161 °C. The current global liquefaction capacity amounts to around 440 tons per day (tpd) allocated to 26 plants, which includes the world's largest plant with a capacity of 90 tpd with three 30 tpd liquefiers commissioned in South Korea in 2024. Currently, only one pilot-scale ship is globally available with the Suiso Frontier having a capacity of 1250 m^3 for a maximum cargo of 89 t H_2 or 3 GWh H_2, whereas it is yet uncertain if high-capacity carriers become operational by 2030. [17–19] The perspective role of LH_2 ship imports is thus still unclear and the potential implementation of large-scale supply chains is expected in the medium to long term, as it is for example stated in Germany's import strategy, whereas all other hydrogen carriers except LOHC are associated with short to medium term implementation periods. [10]

In addition, certification schemes have been developed as a regulatory market framework defining the assessment of the emissions intensity of a clean hydrogen product as a measure of GHG emissions allocated to specific production routes and supply chains. [20] The European Commission defined a GHG emissions intensity threshold of 3.38 kg CO_2(e) per (kg H_2) (28.2 g CO_2(e) per MJ) for renewable fuels of non-biological origin (RFNBO), including renewable hydrogen, as a minimum 70 % GHG emission saving compared to the fossil fuel comparator of 94 g CO_2(e) per MJ in the Renewable Energy Directive (RED) II. [21] The adopted RED III raises the renewable energy targets across all sectors, as 42 % of the industrial hydrogen consumption has to comply with RFNBO criteria in 2030, which is raised to 60 % by 2035 (Article 22a). [22]

2. Literature review

2.1. Studies on hydrogen and PtX imports by ship

Numerous studies explored the techno-economic performance of seaborne supply chains for different PtX products from favourable production sites covering the production and transport costs in near and long-term scenarios. A selection of studies is categorised according to the hydrogen carrier, the CO_2 supply source, the emissions intensity and whether a final delivery of hydrogen is considered by a reconversion of the PtX product at the receiving terminal (see Table 1). Fasihi et al. present long-term potentials of GW-scale methane exports from Patagonia to Japan with supply cost of 66 € per MWh CH_4 based on hybrid PV-wind plants and carbon dioxide (CO_2) supplied from direct air capture (DAC). [23] Another study from Fasihi et al. explored global ammonia production costs based on an hourly electricity supply from PV-wind plants with a cost range of 66–80 € per MWh (345–420 € per t) in 2030 and showed trading options with generic shipping costs. [24] Regarding optimised PtX production costs without transportation, Nayak-Luke and Bañares-Alcántara calculated site-specific levelized cost of ammonia below 59 € per MWh (350 USD per t) in 2030. [25] Salmon and Bañares-Alcántara published an extensive review on ammonia as a prioritized green hydrogen carrier, while a wide disparity of production and transport cost is observed in literature due to unviable and varying plant designs and input data. [26] Niermann et al. explored the efficiency and economics of various LOHC systems releasing the hydrogen stored after shipping with lowest system costs for methanol followed by dibenzyltoluene (DBT). [27] Teichmann et al. showed the hydrogen supply cost for a LOHC ship transport scenario based on N-ethylcarbazole. [27,28] The heat demand for hydrogen release is in both studies either covered by burning a share of hydrogen stored or by external waste heat supply. Schorn et al. calculated long-term methanol import costs at 66–108 € per MWh (370–600 € per t) based on hydrogen production cost of 1.35–2.00 € per kg H_2 and marginal CO_2 supply costs of 100 € per t. [29] A detailed open-source methanol costing model for various scales and locations published by Van Antwerpen et al. shows baseline production cost at 195 € per MWh (1086 € per t) for CO_2 sourced from industrial carbon capture or DAC (1.52 AUD_{2022}/EUR). The long-term potential for LH_2 is given by Heuser et al. with supply cost of 4.40 € per kg H_2 from Patagonia to Japan. [30] Encompassing the four main hydrogen carriers, Hank et al. provide a comprehensive assessment

Table 1
Review of studies focussing on the supply cost and emissions intensity of PtX and hydrogen import scenarios by ship. The studies are categorised in relation to the different PtX products, their CO_2 source and whether a subsequent reconversion into hydrogen is evaluated.

Study	Supply cost of PtX product					Reconversion	CO_2 supply	GHG emissions intensity
H_2 carrier	NH_3	CH_3OH	CH_4	LOHC	Other	(heat source)		
Fasihi et al. [23]			x				DAC	
Fasihi et al. [24]	x						n/a	
Niermann et al. [27]		x		x		x (H_2)	DAC, CCU	
Teichmann et al. [28]				x	LH_2	x (H_2)	n/a	
Schorn et al. [29]		x					DAC, CCU	
Heuser et al. [30]					LH_2		n/a	
Hank et al. [31]	x	x	x	x	LH_2		DAC	
Hank et al. [32]	x	x	x	x	LH_2, FT fuel		DAC, CCU	
Pfennig et al. [33]	x	x	x		FT fuel		DAC	
Ishimoto et al. [34]	x				LH_2	x (H_2)	n/a	x
Lee et al. [35]	x	x		x		x (NG)	Fossil fuel	x
Rezaei et al. [36]	x	x		**x**	LH_2	x (H_2)	DAC, CCU	
IRENA [17]	x			x	LH_2	x	n/a	
Schühle et al. [39]	x	x			DME	x	DAC	
Agora Industrie & TU Hamburg [40]	x	x	**x**	**x**		x (H_2)	DAC, CCU	**x**

of the efficiency and supply costs of PtX exports from Morocco to Germany in 2030 with lowest costs for ammonia (124 € per MWh_{LHV}) compared to LH_2 (126 € per MWh_{LHV}), methanol (131 € per MWh_{LHV}), methane (145 € per MWh_{LHV}) and LOHC (156 € per MWh_{LHV}) based on RE cost of 25 € per MWh and CO_2 sourced from DAC, while a reconversion of the PtX products is not considered. [31] Multiple site-specific PtX pathways are given in an extended study from Hank et al. including a range of supply cost to Germany for LH2, ammonia, methanol and Fischer-Tropsch (FT) derived jet fuels. [32] Global GIS-based PtX production and transport costs in 2050 are calculated by Pfennig et al. based on an hourly temporal and high spatial resolution, which is accessible via the Global PtX Atlas. [33]

In order to meet a dedicated hydrogen demand in the importing country, which is covered by a LOHC supply chain in the mentioned studies, Ishimoto et al. compared the energy efficiency, emissions intensity and hydrogen supply cost for the import of ammonia and LH_2, whereas hydrogen is mostly produced from NG with CCUS at the exporting site. [34] Lee et al. showed the supply cost and emissions intensity for two LOHC systems based on toluene-methylcyclohexane and DBT, ammonia, methanol and LH_2, while hydrogen is produced from fossil energy sources in the base scenario. [35] Rezaei et al. presented total hydrogen supply cost for a delivery of 100 tons per day (tpd) after reconversion for LH_2 (9.36 € per kg H_2), ammonia (6.78 € per kg H_2), LOHC (6.84 € per kg H_2) and methanol (7.74 or 7.80 € per kg H_2 for CO_2 sourced from DAC or CCUS) in an export scenario from Australia to Japan in 2030. [36] A recent report from IRENA evaluates the technological maturity of a LH_2, ammonia and LOHC supply chain including the reconversion step with ammonia leveraging the lowest long-term supply cost, while splitting ammonia back to hydrogen can account to 30–40 % of the total cost. [17] Carbon-based hydrogen carriers like methane, methanol or dimethyl ether (DME) are disregarded in the report as gigatons of CO_2 need to be sourced from DAC or biogenic sources when fossil CO_2 sources are excluded, while both technologies have not been demonstrated at large-scale yet. DAC cost projections remain highly uncertain, as the cost target of 200 € per t CO_2 [37] for commercialized DAC plants at a 1 Million tons per year (Mtpa) scale has been revised by novel cost projections ranging from 272 to 542 USD per t CO_2. [38] Thus, DAC is deemed as an unfeasible supply source for near-term carbon-based PtX products at a GW scale. In this context, Schühle et al. presented a novel DME/CO_2 transport cycle utilizing the higher hydrogen release capacity of DME compared to ammonia and methanol, while CO_2 is captured at the release plant and then back-shipped by the same DME ship thanks to their similar physico-chemical properties forming a closed CO_2 cycle. [39] Since the DME/CO_2 supply chain still faces multiple challenges and industrial-scale projects were not announced yet, DME is not further evaluated in this study focusing on the 2030 timeframe. Following a proposed industrial concept for gigawatt-scale ship imports of synthetic methane to the German port of Wilhelmshaven based on a nearly closed carbon cycle with CCUS, recent studies evaluated the concept of a CO_2/methane cycle in case scenarios for Germany in comparison to a CO_2/methanol transport cycle and other hydrogen carriers. [40,41] The study from Agora Industry and TU Hamburg showed lowest hydrogen supply cost in 2030 for LH_2 (6 € per kg H_2) followed by the methanol cycle (6.20 € per kg H_2), ammonia (7 € per kg H_2), the methane cycle (8 € per kg H_2) and LOHC without external heat supply (9 € per kg H_2), whereas the CO_2 transport cycle effects a cost reduction for the CO_2 supply of 45 % (0.60 € per kg H_2) and 58 % (0.55 € per kg H_2) compared to CO_2 totally sourced from DAC for methane and methanol. [40]

2.2. Limitations in literature and contribution of this study

The review reveals a general gap in literature in terms of calculating hydrogen supply cost after reconverting the carrier for different PtX ship import pathways based on a feasible CO_2 supply with an explicit focus on short to medium term implementation, while evaluating the GHG emissions intensity. To address short-term PtX imports by ship for the ramp-up of a global market in 2030, this study evaluates the economic and environmental feasibility of a CO_2/methane and CO_2/methanol transport cycle based on carbon capture at the reconversion plant in comparison to ammonia and LOHC. These four PtX import routes are identified as the most relevant options to implement near-term hydrogen imports from overseas in a case study for Germany with respect to announced projects, policies and the technology-readiness-level for large-scale deployment. For this purpose, the final hydrogen supply costs are calculated based on a detailed hydrogen carrier and CO_2 transport model. However, it should be noted that these ship-based hydrogen carrier imports need to be compared successively with pipeline-based hydrogen imports or the long-term perspective of LH_2 imports when the PtX supply chain includes a reconversion process to meet a dedicated hydrogen demand. In this context, the perspective of direct hydrogen imports via the European Hydrogen Backbone (EHB) is included here.

To evaluate the climate impact of the CO_2/methane and CO_2/methanol cycle, the carbon footprint of the final hydrogen supply is considered in relation to the GHG emissions at the hydrogen release plant equipped with carbon capture and the CO_2 export by ship. In addition, the emissions intensity assessment incorporates the respective PtX ship fuel consumption for all import routes.

The remainder of this article is structured as follows: The second section presents the scope and scenario design for the four hydrogen supply chains with a detailed techno-economic analysis of all technologies to derive viable parameter assumptions for a GW-scale supply chain scenario in 2030. The third section describes the scope and methodology of the GHG emissions intensity analysis with respect to different marine fuels and the CO_2 cycle. The fourth section presents the results showing the final supply costs and emissions intensity for two PtX import scenarios. The fifth section gives the conclusion and outlook and includes the perspective of emerging opportunities of direct hydrogen import routes over the next decade.

3. Methods and materials

This section derives the techno-economic assumptions of the technologies required for a progressive GW-scale hydrogen export supply chain in 2030 based on four different PtX import routes by ship as shown in Fig. 1. Two potential hydrogen exporters are selected with PtX sites in Australia and Tunisia to evaluate the feasibility of a long-distance and a short-distance shipping route to Germany depending on different electricity costs from VRE as the main cost factor of hydrogen produced from water electrolysis.

3.1. Renewable electricity supply in Australian and Tunisian scenario

The electricity demand of the electrolysis plant and all other facilities is covered by RE supply in the exporting and importing country. A temporal resolution for a site-specific electricity supply from dedicated capacities of solar PV and wind power plants with local capacity factors expressing the full load hours (FLH) of generation at nominal load is not within the scope of this study.

The electricity purchase prices for the electrolysis plants in Australia (AUS) and Tunisia (TUN) refer to country-specific levelized cost of electricity (LCOE) from solar PV as the cheapest electricity source without considering additional costs or grid fees. In general, global weighted average LCOE of utility-scale solar PV projects declined by 89 % between 2010 and 2022 to 49 USD per MWh. [42] Country-specific LCOE for solar PV in Australia are calculated with best estimates of 28 € per MWh in 2021, while future LCOE in 2030 are projected between 17 € per MWh and 36 € per MWh with a minimum capacity factor of 19 % (correspond to 1664 FLH) by CSIRO. [43] An electricity price of 25 € per MWh is assumed for the Australian 2030 scenario. Due to excellent solar irradiation rates in the Middle East and North Africa (MENA) nearly all

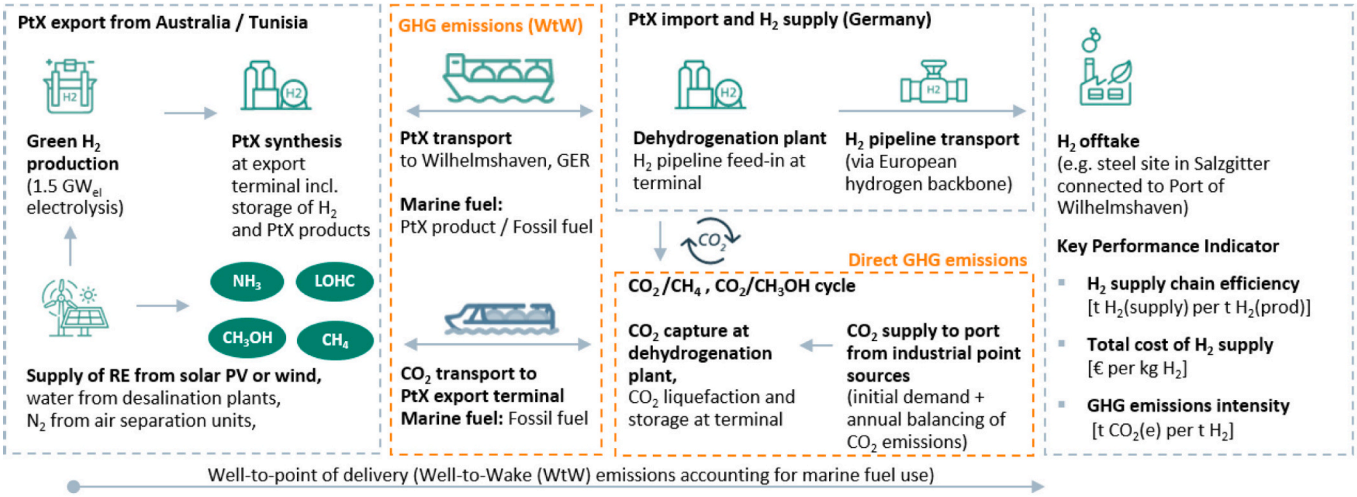


Fig. 1. Scope of renewable hydrogen import routes by ship to Germany for industrial hydrogen offtake in 2030 and system boundary for the assessment of the GHG emissions intensity.

reported bid prices for solar PV projects were below 40 USD per MWh since 2019 with record auction bids of 10.4 [44] and 24.4 [45] USD per MWh in Saudi Arabia and Tunisia, respectively. [46] An electricity price of 30 € per MWh is assumed for the Tunisian scenario to account for the slightly lower average FLH range for solar PV plants in relation to the total share of land area in comparison to Australia (see Table 2). A detailed overview of LCOE comparing onshore wind power to solar PV plants is given in Table 2 with global weighted average LCOE and country-specific estimates for Australia, Tunisia and Germany (GER). The LCOE estimates for onshore wind underline that a combined electricity supply from solar PV and onshore wind would increase the average purchase cost, while guaranteeing higher FLH for the electrolysis plant operation. Wind power capacities provide generally more favourable economic conditions to reduce the hydrogen production cost, whereas the optimum ratio between the electrolysis and renewables capacity strongly impacts site-specific generation cost [50], but can also be optimised over years of operation with added capacities or power purchase agreements. The country-specific LCOE projection for onshore wind in Australia in 2030 with 31 € per MWh on average reveals significantly more optimistic assumptions compared to projections for Germany in 2040 with 52 € per MWh on average. Global weighted average LCOE for onshore wind in 2022 are though reported with 33 USD per MWh, whereas country-specific LCOE vary significantly. [42]

In the baseline scenario for both export sites an electrolysis plant capacity of 1.5 GW is assumed with 3200 FLH requiring an annual RE supply of 4800 GWh. Note, that this leads to an overcapacity of solar PV plants neglecting wind power with for example 2.53 GW of solar PV (1900 FLH) and requires a GW-scale electricity storage system to meet an hourly correlation between generation and electrolysis operation, which is not considered here. To put these renewable capacities into perspective, the total capacity expansion in the MENA region in 2023–2028 is predicted with 60 GW (main case) by the IEA, whereas solar PV contributes 85 % and more than one-third of this expansion is located in Saudi Arabia with around 5 GW dedicated to hydrogen. [51] It should though be noted in view of the two scenarios and assumptions, that Tunisia's installed total RE capacity was less than 0.5 GW with 7 % of the total generation capacity of 6 GW in 2021. In contrast, Australia produced around 31 % of the total generation of 272 TWh in 2022 from RE with a reported solar PV capacity growth of 3.7 GW in 2023. [47,49]

Table 2
LCOE estimates for solar PV and wind power plants for different regions based on average capacity factors.

LCOE [€ per MWh] for	FLH [h per a]	2021–2022	Future	Ref
Solar PV global	1480	46.6	–	[42]
Solar PV AUS	1664-2803	28–41	17–36 (2030)	[43]
Solar PV TUN	1600–1800 [A]	< 40	–	[47]
Solar PV GER	980–1280	31–57	19–35 (2040)	[48]
Onshore Wind global	3241	31.4	–	[42]
Onshore Wind AUS	3066–4030	31–39	25–38 (2030)	[43]
Onshore Wind GER	2500–3200	39–62	34–70 (2040)	[48]

Selected electricity supply cost in this study: 25 € per MWh (Australia), 30 € per MWh (Tunisia). Currency exchange rates: 1.575 AUD_{2021}/EUR, 0.951 EUR_{2022}/USD.

[A] Refers to around 60 % of land area. For Australia nearly 40 % of land area is allocated to a full load hour range of 1800–1900 h per a. [49].

3.2. Hydrogen production from gigawatt-scale water electrolysis

All scenarios consider equal techno-economic assumptions for hydrogen production by operating a proton exchange membrane (PEM) water electrolysis system at a GW scale in 2030. In general, a PEM water electrolyser (PEMWE) is characterized by a high dynamic load range for coupling with intermittent renewable energy sources. [52] Commercially available systems with a 10 MW capacity state a load range of 10–100 % within 30 s, whereas alkaline water electrolysis (AWE) modules of 20 MW report an equal load range. [53,54] State-of-the-art PEM electrolysers were associated with up to 40 % higher capital costs at the system level (700–1400 USD per kW) compared to AWE (500–1000 USD per kW) as the most mature electrolysis technology according to IRENA in 2020. [55] The future cost reduction potential is mainly driven by economies of scale with increasing plant and module sizes and manufacturing scale-up achieving gigawatt production capacities together with specific design improvements. [52,55] Schmidt et al. conducted an expert elicitation study published in 2017 estimating a maximum cost reduction potential through R&D and production scale-up with a mean value of 604 € per kW in 2030 for PEMWE and 500 € per kW for AWE. [56] Capital expenditure (CAPEX) targets announced by the Clean Hydrogen Partnership are 400 € per kW and 500 € per kW for a 100 MW AWE and PEMWE plant in 2030, respectively. [57] IRENA reports capital cost targets below 200 USD per kW for AWE and PEMWE in 2050. [55] In this study, capital cost of 468 € per kW are applied for the 1.5 GW electrolysis plant in 2030 and derived as a mean value from four studies as described in the ESI (ESI-3).

The electricity demand of an electrolysis plant depends on the efficiencies and resulting power consumption at the stack and system level including the electricity demand of balance-of-plant components (e.g. cooling, water treatment, rectifier) and varies during dynamic load

operation. [55] Today's electricity demand of AWE and PEMWE is rated within a broad performance range with best values of 50 kWh per kg H_2 at the system level and 47 kWh per kg H_2 at the stack level, while both technologies have system efficiency targets below 45 kWh per kg H_2 (73 $\%_{LHV}$). [55] Stack degradation increases the electricity demand over the stack lifetime, while the degradation rate target for PEMWE in 2030 accounts to an efficiency loss of 0.12 % when operating 1000 h at nominal load, whereas novel stack designs refer to a loss of 0.09 % per 1000 h. [57,58] The electricity demand of the electrolysis plant in this study is assumed as an average electricity demand over the lifetime and calculated from the initial electricity demand at nominal load with 51.60 kWh per kg H_2 (64.6 $\%_{LHV}$ efficiency) with a constant degradation rate during the entire operation time (0.09 % per 1000 h). Assuming 3200 FLH in the Australian scenario results than in an average electricity demand of 53.10 kWh per kg H_2 over the lifetime (62.8 $\%_{LHV}$). Stack replacement is considered after 80,000 operating hours resulting in an average efficiency of 62.4 $\%_{LHV}$ with a standard estimate for replacement cost with 33 % of the initial electrolyser CAPEX. [52]

The deionized water demand of the plant is assumed with 10 kg H_2O per kg H_2 for PEMWE systems. [53] Water is supplied by a seawater desalination plant based on seawater reverse osmosis (SWRO). Specific capital costs of the SWRO plant are derived from Caldera et al. with 2.23 € per (m^3 a), while further capital costs for an intermediate water storage are considered neglecting water transportation. [59] A conservative electricity demand of 3.75 kWh per m^3 is applied for SWRO based on Hank et al., while an additional electricity demand of 0.45 kWh per m^3 is considered for deionisation based on a process from the ecoinvent v3.3 database. [31,60] Hence, the electricity demand for deionized water supply contributes less than 0.1 % in relation to the electrolysis' electricity demand (0.04 kWh per kg H_2). All assumptions are listed in the ESI (ESI-3).

3.3. Conversion into liquid hydrogen carrier and hydrogen storage

A compressed hydrogen buffer tank storage is applied to guarantee a constant hydrogen supply for a steady-state operation of the PtX synthesis reactor. Dimensioning of the tank storage is simplified in this study since a site-specific and temporal resolution of the electrolysis generation is not considered here. The storage capacity is calculated from the daily average operation time of the electrolyser (8.8 h at full load) for the corresponding hydrogen production rate minus the nominal input of the downstream synthesis reactor. This means that around 60 % of the electrolyser production rate has to be stored at compressed state when the synthesis reactor is operated under steady-state with 8000 FLH, while the storage capacity is sized for 10.5 h of nominal production corresponding to a storage volume of 178 t H_2 with around 14,800 m^3 for a 1.5 GW plant. Capital cost for steel tanks (160 bar) are applied with 500 € per kg. [61] The electricity demand for compression is derived for a tank storage pressure of 120 bar based on the electrolysis output pressure of 30 bar. A viable proof-of-concept of large-scale hydrogen buffer storage systems is missing yet for multi-megawatt scale hydrogen export projects, while other studies assume underground pipe storage or salt cavern systems. [31,32] A detailed site-specific assessment of hydrogen export facilities should therefore include underground hydrogen storage as long-term options. [62,63] All assumptions are stated in the ESI (ESI-3).

Ammonia. Ammonia (NH_3) is a potentially hazardous basic chemical mainly produced as a feedstock for fertilisers and one of the most discussed hydrogen carriers for a global hydrogen trade. A cumulative capacity of 34 Mtpa for projects producing ammonia from green hydrogen was announced by 2021 [17]. The current ammonia production capacity amounts to 243 Mtpa meeting a global demand of around 183 Mtpa in 2020, while today's production is almost entirely sourced from fossil fuels with 72 % based on NG [64]. Market prices typically range between 200 and 300 USD per tonne, while only 10 % of the global demand is traded by ship [64]. With ammonia's high gravimetric hydrogen density of 17.7 % (wt.) around 121 kg of hydrogen can be stored and transported per m^3 of liquid ammonia at a temperature below − 33.3 °C at ambient pressure. Industrial ammonia production is based on the Haber-Bosch (HB) process as an exothermic reaction with a nitrogen and hydrogen feed at temperatures of 350–500 °C and pressures between 90 and 200 bar (Eq. (1)) [65].

$$N_2 + 3\,H_2 \rightleftharpoons 2NH_3,\ \Delta H_R^{STP} = -92.5\,\frac{kJ}{mol} \quad (1)$$

State-of-the art Haber-Bosch reactors with usual capacities of 1000 to 2000 tpd are generally integrated plants based on syngas production from fossil energy sources (e.g. steam methane reforming), while the Haber-Bosch synthesis loop can also be applied or retrofitted for a green ammonia plant coupled to an electrolysis plant. [26,66] One major challenge is the flexibility of the HB reactor enabling a dynamic operation with an integrated flexible-load electrolysis plant. [26] Fasihi et al. report a minimum load of 50 % with a load change limit of 2 % and 20 % for ramp-up and ramp-down, respectively. [24] All synthesis reactors in this study are designed for steady-state operation with 8000 FLH. A high mass conversion rate of 99 % is assumed requiring 0.179 kg H_2 per kg NH_3 and resulting in a hydrogen-to-ammonia conversion efficiency of 86.7 $\%_{LHV}$. [17,24]

A dedicated air separation unit (ASU) is required for the nitrogen supply for green ammonia plants sourcing nitrogen from the air at temperature of −196 °C with a subsequent vaporisation and compression to the reactor pressure. The electricity demand for cryogenic air separation is adopted from a process from Althaus et al. in the ecoinvent database of 0.5–0.8 kWh_{el} per kg_{N2} corresponding to 0.67 kWh_{el} per kg NH_3 for the upper value. [67] Additional electricity is needed for nitrogen and H_2 feed compression and the recycle loop with 0.48 kWh_{el} per kg_{NH3}. [31] The total electricity demand is then assumed with 1.15 kWh_{el} per kg NH_3 as a pessimistic value while other studies report a consumption down to 0.57 kWh_{el} per kg_{NH3}. [17]

The applied capital cost for the entire ammonia plant including the ASU in this study is derived from the cost calculation by Fasihi et al. in relation to Morgan with 575 € per (t NH_3 a) for a production capacity of 500,000 tpa corresponding to 890 € per $kW_{NH3,LHV}$ for 8000 FLH. [24,68] Fasihi et al. shows that the effect of economies of scale is negligible for capacities larger than 250,000 tpa based on Morgan. [24] Other studies assume a more progressive cost reduction with 500 € per $kW_{NH3,LHV}$ for a capacity of 1500 tpd with a further reduction potential for larger capacities as outlined by IRENA. [17] The H21 North of England report estimates for instance total project cost of £2473 M for a 5000 tpd facility coupled to an electrolysis plant and export infrastructure with a jetty, tanks and a loading facility including owner's cost. [65] This corresponds to an estimated capital cost of around 595 € per (t NH_3 a) or 980 € per kW (NH_3 LHV) for the ammonia facility after deducting the electrolyser capital cost of £1160 per kW (H_2 HHV) (conversion rate 1.13 €/£). [65] The techno-economic assumptions for all PtX synthesis reactions are stated in Table 3.

Methanol. Methanol (CH_3OH) is the simplest alcohol with a boiling point at 65 °C and a density of 791 kg per cubic meter (20 °C) with a gravimetric hydrogen storage capacity of 12.5 % (wt.) corresponding to a volumetric hydrogen storage density of 98.9 kg H_2 per m^3 [69]. Global demand for methanol is forecasted by IRENA to reach 120 Mt in 2025 from 98 Mt in 2019, while only 0.2 Mt of today's production is based on renewable sources [70]. Current production costs for methanol produced from fossil fuels (NG or coal) range between 150 and 200 USD per t CH_3OH [70]. Renewable methanol can either be produced from biomass feedstocks (bio-methanol) or synthesized from green hydrogen and CO_2 sourced from direct air capture (DAC) or from bioenergy with carbon capture and storage (BECCS), also referred to as e-methanol [70]. In this study, e-methanol is assessed as a hydrogen carrier with a subsequent reconversion at the import terminal to meet a dedicated green hydrogen demand, while CO_2 is sourced from industrial point sources. The production of e-methanol from green hydrogen and CO_2 has already

Table 3
Techno-economic assumptions for hydrogen carrier synthesis and reconversion plants.

Parameter		NH_3		CH_3OH		CH_4		LOHC	
		Synthesis	Reconversion	Synthesis	Reconversion	Synthesis	Reconversion	Synthesis	Reconversion
Conversion efficiency	$\%_{LHV}$	86.7	90.9	86.9	87.4	82.90	79.89	99.0	65.7
Conversion rate	kg H_2 per kg PtX	0.179	0.141	0.191	0.145	0.503	0.324	–	–
CAPEX	€ per kW	890	331	235	844	295	844	260	261
Electricity	kWh per kg	1.15	0.47	0.31	1.91	0.67	1.91	0.37	0.37

PtX or hydrogen production plant capacities in kW are based on the LHV of the respective energy carrier. Further assumptions and the reference plants are stated in the ESI (ESI S-3).

been demonstrated in several plants, while the "George Olah" plant in Iceland is recognized as the first commercial plant commissioned in 2011 with an annual production volume of 4000 tons [70]. New renewable methanol projects are announced with project production capacities up to 180,000 tpa in different countries (e.g. in Australia, Sweden or Canada) [70].

The catalytic synthesis of methanol is achieved in reactors at a temperature of 220–280 °C and a pressure of 10–80 bar, while three reactions (Eq. (2), (3), (4)) are taking place depending on the reaction kinetics and conditions. [29,69] The equilibrium reaction of the direct hydrogenation of CO_2 requires a feed of 0.189 kg H_2 per kg CH_3OH and 1.374 kg CO_2 per kg CH_3OH for one kilogram of pure methanol obtained after water separation through distillation. [29] The electricity demand of the synthesis process including compressors for the CO_2 and H_2 feed is assumed with 0.31 kWh per kg CH_3OH based on a process modelling by Hank et al. producing 36,500 tpa with a given hydrogen conversion rate of 0.20 kg H_2 per kg CH_3OH. [31] This conversion rate corresponds to a hydrogen-to-methanol conversion efficiency of 82.9 $\%_{LHV}$ compared to the ideal efficiency of 87.9 $\%_{LHV}$. Van Antwerpen et al. report a methanol yield of 95 % as a ratio between the methanol production and the CO_2 feed flow rates due to CO_2 and methanol vapour losses occurring in the purge and distillation units, while a total hydrogen-to-methanol conversion rate of 0.210 kg H_2 per kg CH_3OH is achieved. [71] An Australian project from ABEL Energy with a production capacity of 226 tons per day based on a 100 MW electrolysis plant refers to a hydrogen conversion rate of 0.191 kg H_2 per kg CH_3OH (86.9 $\%_{LHV}$) based on the production rates given in the knowledge sharing report, which is applied here as an near-ideal reaction. [72] The total electricity demand for the hydrogen and subsequent methanol production accounts to 10.33 kWh per kg CH_3OH.

Direct hydrogenation of CO_2 : $CO_2 + 3\,H_2 \rightleftharpoons CH_3OH + H_2O, \Delta H_R^{STP}$

$$= -49.5\frac{kJ}{mol} \quad (2)$$

Reverse water gas shift reaction : $CO_2 + H_2 \rightleftharpoons CO + H_2O, \Delta H_R^{STP}$

$$= 41.2\frac{kJ}{mol} \quad (3)$$

Hydrogenation of CO : $CO + 2H_2 \rightleftharpoons CH_3OH, \Delta H_R^{STP} = -90.7\frac{kJ}{mol}$ (4)

The methanol synthesis plants in the scenarios exceed a capacity of 300 MW(CH_3OH). The capital cost can only be derived from component cost estimations and economies of scale for different plant capacities. Brynolf et al. report a cost range of 200–400 € per kW CH_3OH for a 200 MW(CH_3OH) plant, while Schorn et al. assume capital cost of 235 € per kW(CH_3OH) for a 300 MW(CH_3OH) capacity derived from the plant design that is presented by Schemme et al. [29,73,74] The effect of further decreasing specific capital cost is limited for plants sizes larger than 200 MW(CH_3OH) as reported by Schemme et al. [74] Thus, capital cost of 235 € per kW(CH_3OH) are assumed as a lower limit for all scenarios here.

Methane. Converting green hydrogen into liquified synthetic methane offers a significant higher volumetric energy density (21.1 MJ per m^3) compared to LH_2 (8.5 MJ per m^3) with a low boiling point at −161.6 °C (1 bar) corresponding to a volumetric hydrogen storage density of 106.1 kg H_2 per m^3 with a high gravimetric hydrogen storage capacity of 25.1 % (wt.).

The Sabatier reaction describes the conversion of green hydrogen and CO_2 into synthetic methane (CH_4), also referred to as e-methane, e-NG or synthetic natural gas (SNG), according to the CO_2 hydrogenation reaction (Eq. (5)) as a catalytic methanation process and a highly exothermic reaction.

Sabatier reaction : $CO_2 + 4\,H_2 \rightleftharpoons CH_4 + 2\,H_2O, \Delta H_R^{STP} = -165\frac{kJ}{mol}$ (5)

A CO_2 conversion rate of 99 % correlates to a methane content of 95 % in the e-NG product gas. Different reactor concepts are applicable at an operation temperature from 250 to 550 °C and a pressure range of 1–100 bar, while a more advanced temperature control is needed compared to less exothermic synthesis reactions like methanol. High temperature heat is released and additional heat can be sourced from methane product gas cooling. The excess heat can then be used for steam and power generation to supply additional electricity to an integrated electrolysis and sabatier plant. In addition, around 4.5 kg of water are produced per kg of hydrogen feed, so that a circular water cycle could reduce the total freshwater demand for hydrogen generation by 45 %. [75,76]

The load change rate and the minimum load determine the process flexibility for a dynamic system operation to minimize the need for hydrogen buffer storage. [75] The theoretical maximum efficiency of a Sabatier reactor with a conversion rate of 100 % producing 2.0 kg of methane from 1.0 kg of hydrogen is 77.85 $\%_{LHV}$ (or 82.95 $\%_{HHV}$). The methanation process in this study assumes an ideal 100 % conversion rate requiring 0.503 kg H_2 per kg CH_4 as applied by Gorre et al. [77] (LHV of 50 MJ per kg CH_4). Hank et al. apply a conversion rate of 0.52 kg H_2 per kg CH_4. [31] Demonstration plants like the STORE&GO Falkenhagen power-to-gas plant in Germany with a capacity of 0.58 MW(CH_4) achieved a methane fraction of more than 99 $\%_{vol}$ in the product gas and load changes between 40 and 100 % with an innovative honeycomb reactor. [78]

Reliable capital cost data for multi MW-scale sabatier reactors integrated into PtX plants is scarce since only a few demonstration plants have been deployed. [79] The Audi e-Gas plant in Werlte commissioned in 2013 is recognized as the largest power-to-gas plant worldwide with a capacity of 3.2 MW_{CH4}. [79,80] Götz et al. report high uncertainty issues in capital cost estimates for catalytic methanation and refer to an industrial reference of 130 $€_{2014}$ per kW(CH_4) with economies of scale for a 110 MW plant from Outotec GmbH. [75] Gorre et al. state capital cost of 600 $€_{2017}$ per kW(CH_4) for a small-scale system coupled to a 1 MW electrolyser in 2017 and future cost of 295 € per kW(CH_4) (2030 with 50 MW electrolysis) and 195 € per kW(CH_4) (2050 with 50 MW electrolysis) with reference to the STORE&GO project report. [77,79] The demonstration plant in Falkenhagen reports capital cost of 3740 € per kW(CH_4) with a reduction potential to 360 € per kW(CH_4) for a 50 MW plant size, while further learning effects can leverage a cost range of 120–290 € per kW by 2050. [78] In this study capital cost of 295 € per kW_{CH4} are assumed. The electricity demand of the methanation plant for the

operation of pumps and compressors is assumed with 25 kW_{el} per MW_{el} of electrolysis capacity for steady-state operation corresponding to an electricity consumption of 0.67 MWh per t CH_4. [77]

The e-NG product is directly liquified and stored in tanks at the export terminal. Additional transport demand is neglected. Standard liquefaction plant capacities for LNG exports range between 0.5 and 8 Mtpa. [81] The export volumes here account to 0.180 Mtpa. A typical natural gas liquefaction facility includes different gas processing steps such as gas sweetening or dehydration which are not included here in the cost assessment and a liquefaction train with heat exchangers, compressors, expanders and gas turbines. [82] Zhang et al. gives a capital cost.

range from 125 to 2255 USD per tpa, while most liquefaction trains are reported with capital costs below.

1000 USD per tpa. [83] Raj et al. calculate a total CAPEX of 1200 USD per tpa for a 10 Mtpa facility with three storage tanks of 160,000 m^3 in Canada. [82] In this study, a CAPEX of 620 USD per tpa is assumed for the liquefaction system of e-NG based on an industrial cost reference from Linde Group for a single mixed refrigerant cycle with a capacity of 0.2 Mtpa. [84] Hank et al. apply similar capital costs for synthetic methane with 604 USD per tpa. [31] The electricity demand for liquefaction is considered with 0.5 kWh per kg CH_4 as a common base assumption. [31]

Liquid Organic Hydrogen Carriers. LOHC comprise various chemical substances (e.g. hydrocarbons) for storing hydrogen in a chemical compound (hydrogenation) and releasing hydrogen (dehydrogenation) at the point of demand, whereas the unloaded LOHC substance is transported back to the hydrogen production site to be reloaded with hydrogen for the next transport cycle. [28] One key parameter in terms of economics and sustainability of a LOHC-based supply chain is the heat demand for dehydrogenation varying for different LOHC compounds and equalling the amount of released heat during hydrogenation. [17] The German company Hydrogenious LOHC Technologies applies benzyltoluene (BT) in their LOHC system, which is commonly used as a heat transfer oil. [17,85] Announced industrial scale projects from Hydrogenious like “Northern Green Crane” outline (de-)hydrogenation plant capacities of 12 tpd with a hydrogen import volume of 8000 tpa. [85] Rüde et al. show multiple benefits for benzyltoluene (BT) as a highly promising LOHC system (benzyltoluene (H0-BT)/perhydro benzyltoluene (H12-BT) compared to dibenzyltoluene and toluene as other discussed substances. [86] In this study, BT is applied in the LOHC system with a gravimetric hydrogen storage density of 6.22 wt% corresponding to 54.5 kg H_2 per m^3 with a density of 876 kg per m^3 (H12-BT, at 20 °C) and 996 kg per m^3 (H0-BT, at 20 °C). [86] The purchase cost for BT are assumed with 1.60 € per kg, while the BT production costs mainly depend on the toluene feedstock price and are estimated within a range of 1.42 to 1.80 € per kg for a 100,000 tpa production capacity and a toluene price ranging from 600 to 900 € per t. [87]

The initial BT demand of around 0.42 Mio. tpa in the Australian LOHC(BT) scenario in this study also indicates that a dedicated production scale up of BT is needed since the current market for all heat transfer fluids accounts to around 0.5 Mio. tpa, while there is no raw material scarcity predicted related to the required toluene and chlorine feedstock volumes. [87] The initial BT demand in the Tunisian scenario with one operating LOHC(BT) ship is based on the required BT feed for dehydrogenation during one ship roundtrip times a factor of 1.20 and the onboard stored BT volume for one ship. The BT demand in the Australian scenario with three required LOHC(BT) ships allocates only to the onboard stored BT volumes. Annual replacement costs of BT due to degradation and cycle losses are assumed with 0.1 wt%, while catalyst replacement costs are not considered. [88,89] The exothermic hydrogenation process is carried out at medium temperature (150–250 °C) and elevated pressure (e.g. 30 bar for BT). [17,86] The electricity consumption basically for hydrogen compression is assumed with 0.37 kWh per kg H_2. [28]

Future capital costs are reported by IRENA within a range of 60 to 120 USD per kW(H_2) for large-scale plants (up to 1000 tpd) comprising literature estimates based on scaling factors, while plant scale-up is not associated with major barriers. [17] CAPEX for the hydrogenation unit are applied with 260 € per kW(H_2) assuming a cost reduction potential through economies of scale for a larger plant capacity in this case study with a common scaling factor of 0.60 compared to the reference plant design from Hank et al. based on data from Hydrogenious (see Eq. (A.3) in ESI-3) [31]. Details about the techno-economic assumptions for all PtX and reconversion plants and reference plant data are given in the ESI (ESI-3).

Dehydrogenation as an endothermic catalytic reaction is processed with a platinum-based catalyst at medium temperatures above the boiling point of the LOHC compound (e.g. 290 °C for BT) releasing hydrogen at low pressure (e.g. 1.6 bar) and thus requiring downstream hydrogen compression. [17,86] The required heat for dehydrogenation for H12-BT is based on the specific enthalpy of reaction with 63.5 kJ per mol H_2 (31.5 MJ per kg H_2 or 8.75 kWh per kg H_2) [90], which is covered by the combustion of hydrogen released from the LOHC system applying an efficiency of 80 $\%_{LHV}$ for hydrogen burners and 97.5 $\%_{LHV}$ for heat exchangers as assumed in [91]. This results into an internal energy consumption of 11.22 kWh per kg H2 and a hydrogen loss of 33.7 %LHV in relation to the theoretical hydrogen release rate (as shown in the ESI-3 Eq. (A-4)). The total hydrogen release efficiency accounts then to 65.7 %LHV considering 1 % of additional hydrogen losses in the reactor. [98] The total efficiency accounts then to 65.7 $\%_{LHV}$. Eypasch et al. calculate for instance with a higher energy consumption of 12.54 kWh per kg H_2 based on an enthalpy of reaction of 71 kJ per mol H_2 [91], while Hydrogenious states a heat demand of approximately 11 kWh per kg H_2 [99]. Hank et al. calculate with a reduced hydrogen consumption of 25 % (8.3 kWh per kg H_2) and 30 % (10.0 kWh per kg H_2). [31] Site-specific potentials of integrating waste heat from industrial sites at the port can avoid hydrogen burning and thus increase the total economic performance, which is presented in in a JRC study with heat supply cost at 20 €/MWh for a heat demand of 11.4 kWh per kg H_2. [89] CAPEX for the dehydrogenation plant are derived with 261 € per kW(H_2) applying a scaling factor of 0.60 for the reference plant given in [31] to account for an indicated cost reduction potential for larger plant capacities with lower capital cost of 100 to 250 USD per kW(H_2) as outlined by IRENA [17,31].

Storage at export terminal. Intermediate tank storage of the liquid PtX product is considered at the export terminal with a storage capacity dimensioned in relation to the PtX production during one roundtrip ship voyage times a factor of 1.20. The storage capacities at the import terminal are based on the respective single PtX ship capacities. For the LOHC(BT) supply chain the export and import storage is increased by the size of one ship cargo capacity, respectively, for a separated storage of H12-BT and H0-BT.

3.4. Shipping and terminal cost model

The sea transport of the liquid hydrogen carrier refers to standard carriers in the LNG, methanol and ammonia market. Common LNG tanker capacities of 175,000 m^3 are utilized for e-NG, while more than 600 LNG tankers were globally operational in 2022. [81] The handling of LOHC is similar to oil products, which enables the use of high capacity tankers for oil and chemicals from the Suezmax ship class. LOHC(BT) tankers are considered with a cargo capacity of 150,000 t. [17] Ammonia is shipped with LPG tankers at capacities up to 40,000 t, while a fleet of 170 operating vessels is available today from around 1000 vessels available for LPG. [100,101] A capacity of 51,000 t is considered, while new ship orders for capacities of 62,000 t with a volume of 93,000 m^3 were reported. [102] A methanol vessel capacity of 43,000 t is applied based on common ocean-going vessels. [94] It is assumed that ammonia, methanol and methane engines are applied as propulsion systems in the PtX-fuelled scenarios for the respective vessels. The

exported PtX product is then used as a marine fuel reducing the PtX import volume. Boil-off gas (BOG) is included in the methane fuel consumption. First vessel demonstrations with ammonia two-stroke engines are expected in 2024. [103] The fuel consumption is adopted from conventional engine efficiencies here. Methanol dual-fuel engines have already entered the market, while 22 methanol-fuelled vessels were operating by February 2023. [104] In contrast, over 900 LNG-fuelled vessels were available in 2022. [104] The LOHC carriers are based on a conventional propulsion for large tankers running on heavy fuel oil (HFO), as an onboard dehydrogenation-based propulsion is far away from commercialization especially for large carriers and long distances. [17] The CO_2 transport ships consume likewise HFO.

The calculation of the shipping cost is based on a charter rate cost model for a scenario-specific fleet dependent on the distance and export volume. The daily charter rate is a function of the capital cost of the vessel, operation and maintenance cost, crew cost and fleet utilization rate. The OPEX include a Suez Canal fee for the Australian route, a port fee and insurance cost adopted from the open-source model for hydrogen shipping cost from Johnston et al. [105] The main assumptions for the shipping cost model are stated in Table 4.

The e-NG import considers an onshore regasification unit with a capacity of 3 Mtpa e-NG or LNG. This includes several onshore terminal units (e.g. a jetty with piping and unloading lines) separate from tank capacities with a total CAPEX of USD 620,000,000. [106] The terminal costs are proportionally calculated considering a terminal utilization of around 6 % for the PtX route. A full utilization would require an electrolysis capacity of around 25 GW at 3200 FLH or 10 GW at 8000 FLH equivalent to 41.6 TWh_{LHV} of synthetic methane.

3.5. Reconversion of hydrogen carrier

Ammonia cracking. Ammonia can be reconverted to hydrogen at high temperatures, which is called ammonia cracking or splitting according to Eq.(6) [69].

$$2NH_3 \rightleftharpoons 3\,H_2 + N_2,\ \Delta H_R^{298K} = 46.2\frac{kJ}{mol} \tag{6}$$

The spontaneous decomposition of ammonia starts at temperatures above 200 °C, while different catalysts are applicable to achieve a complete conversion of the ammonia feed. [17,69] Conventional catalysts based on nickel or cobalt require higher reaction temperatures (600–900 °C), while more expensive and more active catalysts like ruthenium enable lower temperatures (350–600 °C). [17,69] Lower reaction pressures at ambient conditions are favoured to achieve a high conversion rate and low ammonia slippage but increase the subsequent energy demand for hydrogen compression. [17] Thus, industrial plants like in the H21 project are designed at higher pressures (20–40 bar, 900 °C) with a daily conversion of around 6500 t NH_3 producing 1.5 GW_{HHV} of hydrogen and a feed conversion rate of 99.5 %. [65] The process design requires an efficient heat recovery for integrated preheating to decrease the energy consumption and a subsequent purification of the cracked ammonia product containing around 75 % hydrogen to deliver pure hydrogen. [17,65] The total energy consumption of the plant depends on the energy demand for preheating, the cracking process and additional steam generation and occurring heat losses in the process. [17] Assumptions for the energy consumption that is internally covered from the hydrogen-ammonia feed range between 15 $\%_{LHV}$ and 30 $\%_{LHV}$ of the produced hydrogen energy due to different system boundaries and process designs in literature. [17] In this study, the energy consumption is derived from the ammonia cracking plant in the H21 project, which is given with a high conversion efficiency of 90.86 $\%_{LHV}$ (88.81 $\%_{HHV}$) excluding the electricity demand corresponding to a hydrogen yield of 0.141 kg H_2 per kg NH_3 and a respective hydrogen loss of 20.4 % related to hydrogen stored as ammonia. [65] Note, that this cracking plant efficiency can be reviewed as an optimistic performance assumption compared to other studies as outlined by IRENA and related to base assumptions from industrial companies that 25 % of hydrogen stored in ammonia is lost in the process corresponding to a conversion efficiency of 85.6 $\%_{LHV}$. [17] Some studies assume even lower conversion efficiencies like Ishimoto et al. with a hydrogen yield of 0.123 kg H_2 per kg NH_3 (79.6 $\%_{LHV}$) after purification and hydrogen consumption for cracking. [34] The electricity demand of 0.47 kWh per kg H_2 based on the H21 plant design accounts for a hydrogen export pressure of 80 bar. [65] The energy efficiency of the plant amounts then to 89.7 $\%_{LHV}$.

Large-scale ammonia cracking plants converting up to 2400 t NH_3 per day have only been used for heavy water production for nuclear reactors without purification for a dedicated pure hydrogen demand. [17,65] The process design is though similar to steam methane reforming (SMR) plants to derive cost estimates from commercialized SMR plants. [17,65,107] The here applied capital cost of 331 € per kW (H_2 LHV) is derived from the total project cost in the H21 report with £372,000,000 including contingency and owner's cost for the import terminal (average conversion rate 0.885 $£_{2018}/€$). [65] This assumption coincides with many literature estimates showing capital cost below 500 € per kW(H_2) for plant capacities of 1000 t NH_3 per day when further cost reductions through economies of scale are negligible with capital cost of around 250 € per kW(H_2). [17] The ammonia cracker in the case

Table 4
Techno-economic assumptions for ship transport of hydrogen carrier and CO_2.

Ship parameter		NH_3	CH_3OH	CH_4	LOHC(BT)	CO_2
Cargo capacity (98 % filling of capacity)	m^3	74,450	54,223[a]	171,500	172,345	50,000
	t	50,771	42,891	72,545	150,974[b]	57,500
	GWh_{LHV}	267	239	1008	312	-
Fuel consumption	$MWh_{HHV}\ km^{-1}$	1.09 [c]	0.92 [c]	2.00[c]	2.52[d]	0.67 [92]
(distance / loaded cargo)	kWh_{HHV} $(t\ 1000\ km)^{-1}$	21.5	21.4	27.6	16.7	11.6
Speed	$km\ h^{-1}$	29	26	33	26	26
CAPEX	€	62,377,060 [e]	44,462,911 [f]	201,710,000 [g]	141,321,824 [e]	80,900,000 [93]
Charter rate	€ day^{-1}	30,061	22,608	91,098	62,117	39,385

[a] Reference carrier: Taranaki Sun.
[b] LOHC H12-BT cargo capacity based on H12-BT density of 876 kg m^{-3}.
[c] Fuel consumption based on data from GMW Consultancy for the applied ship classes in the "GreenH2SZ" project and corresponding to similar carriers applied in [94].
[d] Data from GMW Consultancy for a crude tanker with a maximum cargo capacity of 180,000 t.
[e] Based on average specific CAPEX of 863 USD/m^3 for LPG carriers (82,000 m^3 capacity class). [95].
[f] Based on average specific CAPEX of 845 USD/m^3 for crude oil carriers (neglecting the different ship classes for LOHC(BT) and methanol). [94].
[g] Based on average specific CAPEX of 1322 USD/m^3 for LNG carriers derived from an end of 2022 order estimate. [81] HFO cost with 465 € per t based on the HFO global average bunker price (IFO380) with 488.50 USD per t (06/2021–06/2023). [96] Charter rate based on CAPEX and own calculation model. Calculation model and further assumptions are given in the ESI (ESI-3). Average Suez Canal fee of 400,000 USD assumed for all carriers (0.877 EUR_{2020}/USD). [97].

scenarios accounts to a capacity of 1375 t NH_3 per day.

Methane reforming. For the reconversion of synthetic methane today's state-of-the art technologies for hydrogen production from natural gas can be applied: SMR or autothermal reforming (ATR). Both technologies require temperatures of more than 900 °C to produce a synthesis gas of H_2, carbon monoxide and CO_2 from the steam and methane feed in the reactor at high pressures. For SMR the required heat is supplied to the reformer by an external combustion of a fuel gas, which consumes around 35 % of the natural gas feed, while for ATR heat is generated inside the ATR reactor by partial combustion of the reformer feed with oxygen. Thus, ATR yields a more CO_2 rich synthesis gas, while the external combustion for SMR generates a separate flue gas containing CO_2, which increases the costs for carbon capture. The produced synthesis gas in the reformer unit is then reacted with water to produce additional H_2 and CO_2 from the CO fraction in the water-gas shift section with the overall reaction shown in Eq. (7).

$$CH_4 + H_2O + \frac{1}{2}O_2 \rightleftharpoons CO_2 + 3H_2, \Delta H_R^{298K} = -22.8\frac{kJ}{mol} \quad (7)$$

Therefore, hydrogen produced from a synthetic methane feed is obtained from both methane and water. The hydrogen-rich synthesis gas mainly containing H_2 and CO_2 is then cleaned in a following pressure swing adsorption system to deliver the required hydrogen purity of more than 99 % by removing CO_2, impurities (e.g. carbon monoxide) and unreacted methane [65,108]. In this study, an ATR-CCUS plant is applied with a 94.1 % carbon capture rate based on the proposed process design for natural gas reforming from the H21 project report. This ATR process applies a syngas high pressure CO_2 capture technology. The conversion efficiency of the plant is reported with 79.89 $\%_{LHV}$ (83.10 $\%_{HHV}$) excluding the electricity demand corresponding to a hydrogen yield of 0.324 kg H_2 per kg CH_4. The energy demand is covered by consuming the renewable methane feed. The specific electricity demand is assumed with 1.91 kWh per kg H_2 based on the H21 plant design with a hydrogen export of 1.5 GW_{HHV} at 80 bar resulting in an overall energy efficiency of 74.54 $\%_{LHV}$ (79.89 $\%_{HHV}$). [65] As an example for novel plant designs with even higher carbon capture rates, Antonini et al. present a techno-environmental analysis for a vacuum pressure swing adsorption with ATR and SMR as a novel CCUS technology with a 98 % capture rate, while the energy conversion efficiencies are lower for an ATR plant configuration with 76.2 $\%_{LHV}$. [108] In contrast to ammonia cracking, SMR and ATR are widely deployed for fossil-based hydrogen production, whereas only 16 hydrogen plants are in general equipped with CCUS capturing around 11 Mtpa based on capture rates below 60 % for natural gas reforming. [4] Two ATR plants with capture rates of 90 %–95 % are currently being built in North America. [4] The here applied capital cost of 844 € per kW(H_2) refer to the total project cost of £947,000,000 (average conversion rate 0.885 $£_{2018}/€$) including engineering, procurement and construction and other costs in the H21 report based on a KBR cost reference and an internal calculation from Equinor. [65] The ATR plant here accounts to a production capacity of 220 MW_{LHV}.

Methanol reforming. Converting methanol to hydrogen can be processed by methanol steam reforming (MSR) according to Eq. (8) at lower reactor temperatures (250–300 °C) compared to methane reforming with a subsequent pressure swing adsorption (PSA) unit for producing pure hydrogen from synthesis gas. [109,110]

$$CH_3OH + H_2O \rightleftharpoons CO_2 + 3H_2, \Delta H_R^{298K} = -49.5\frac{kJ}{mol} \quad (8)$$

The theoretical hydrogen release from MSR accounts to 18.8 wt% in relation to the methanol feed under ideal conditions, which is higher than the hydrogen storage capacity of methanol with 12.5 wt%, since one extra mol of hydrogen is provided from water per mol of methanol during MSR. Thus, the theoretical hydrogen release from MSR is also higher than for ammonia cracking with 17.8 wt%. [39]

In contrast to ammonia, e-methanol is rarely discussed as a hydrogen transport carrier with a subsequent reforming process, and rather foreseen to meet a growing methanol demand. [70] The techno-economic assumptions for a large-scale MSR in this study cannot be derived from a technical proof-of-concept like for ATR-CCUS and ammonia cracking. MSR plants are only commercially available at a small-scale up to a hydrogen production of around 11 tpd or 15 MW (H_2) like plants from Mahler AGS with a hydrogen conversion rate of 0.143 kg H_2 per kg CH_3OH (86.0 $\%_{LHV}$). [111] The here applied energy consumption for MSR is based on the heat demand for MSR of 1.14 MWh per kg methanol feed given by Schühle et al., whereas more than 50 % of this energy is needed for heating and evaporation of the feed. [39] This results into a heat demand of 6.04 kWh per kg H_2 released based on the theoretical hydrogen release of 18.8 wt%. Andersson et al. report a heat demand of 6.70 kWh per kg H_2. [69] Similarly to the heat supply for LOHC dehydrogenation, a combustion of the product gas of the reformer results into an energy consumption of 7.74 kWh per kg H_2 released (as calculated in the ESI-3 Eq. (A-4)).

A MSR conversion efficiency of 0.145 kg H_2 per kg CH_3OH (87.4 $\%_{LHV}$) is thereby applied here. This corresponds to a hydrogen loss of 23 wt% in relation to theoretical hydrogen product released from methanol. This theoretical assumption for the MSR energy consumption neglecting the energy demand for a CO_2 capture technology is slightly higher than the given efficiency of a small-scale reference plant. A more reliable process modelling of the MSR plant requires a detailed feasibility study with the integration of a synthesis gas or flue gas CO_2 capture technology as shown in the H21 report. The specific electricity demand is adopted from the ATR process resulting into a total energy efficiency of 83.2 $\%_{LHV}$.

The capital cost of MSR with CO_2 capture is adopted from the ATR plant assumptions with 844 € per kW(H_2) as a conservative cost assumption, although MSR is associated with a relatively low capital cost potential compared to methane reforming and ammonia cracking. The only reference cost for a dedicated MSR plant without considering CO_2 capture is presented by Andersson with approximated specific capital cost of 120 € per kW(H_2) (around 50 M€ for a capacity of 300 t per day) in relation to a white paper from Lurgi in 2001. [109] The MSR-CCUS plant in this study accounts to a hydrogen production capacity of 250 MW_{LHV}.

3.6. Pipeline transport to industrial offtaker

A hydrogen pipeline transmission over a distance of around 300 km is assumed as an exemplary scenario for a final delivery from Wilhelmshaven to the steel production site in Salzgitter and based on cost assumptions of the European Hydrogen Backbone for a newly built transmission pipeline (4.1 GW_{LHV}) [12]. The assessment of a perspective direct pipeline transmission from Tunisia to Germany as proposed in the North Africa Southern Europe corridor over a distance of around 4000 km to Salzgitter is based on recent cost assumptions for the EHB, which are given in the ESI (ESI-3) [13].

3.7. Carbon dioxide transport cycle

The CO_2 transport cycle considers dedicated liquid CO_2 carriers for the export from the import terminal in Germany back to the hydrogen export site to meet the annual CO_2 demand for the methanol and methane route. The annual CO_2 export volume is thereby defined by the specific CO_2 demand of the installed PtX synthesis capacity in the scenarios and primarily met by the CO_2 capture at the reforming plants. The uncaptured CO_2 emissions are offset by an additional CO_2 supply from external sources delivered to the PtX import terminal. Further potential occurring CO_2 losses during storage and shipping (e.g. BOG losses) are not considered, so that a closed carbon loop is assumed for the captured CO_2 volume. The CO_2 feed from carbon capture is directly liquified and then stored at the export terminal before shipping. The CO_2 supply costs at the hydrogen export site include the costs for liquefaction, buffer

storage with loading facilities at both terminals and shipping according to the shipping cost model for the hydrogen carrier transport. Supply cost of 50 € per t CO_2 are assumed for the initial and annual CO_2 demand.

The techno-economic assumptions for the liquefaction step of a pure CO_2 feed at ambient conditions are derived from Roussanaly et al. for a large-scale CO_2 liquefaction plant (1 Mtpa) based on the process design of Deng et al. (see ESI-3). [93,112] This includes the electricity demand for compression of a pure CO_2 feed to the transport pressure neglecting major impurities. [93] The terminal cost include loading and unloading facilities. [93] The export terminal storage capacity considers the annual export volume and roundtrip voyage time of one CO_2 ship. The import terminal storage capacity is dimensioned related to the ship capacity times a dimensioning factor of 1.2. [93] The techno-economic ship design assumptions are adopted from Roussanaly et al. for a future CO_2 carrier capacity of 50,000 t CO_2 at a transport pressure of 7 bar (−46 °C) near triple point conditions as a more cost-effective transport pressure compared to 15 barg shipping. [93] Today's ship transport of food-grade CO_2 is established with 15 barg and maximum tanker capacities of 1800 kt CO_2. [92,93] First ground-breaking ship developments are announced with capacities of around 8000 t CO_2 in the Norwegian "Northern Lights" project. [113] It is noteworthy that the annual CO_2 demand in the scenarios requires CO_2 export volumes of around 500,000 tpa. A potential deployment of dual-use e-NG/CO_2 carriers that are technically capable to transport liquid CO_2 in modified double-hull storage tanks is not considered in this study. [114] Such dual-use ships could significantly reduce the transport costs in the proposed CH_4/CO_2 cycle, as the CH_4 carrier avoids an empty return carrying captured CO_2 on the voyage back to the production site.

3.8. Final hydrogen supply cost calculation

The economic performance of the scenarios relates to the amount of hydrogen supplied to an industrial site in Germany. Thus, the total cost of hydrogen delivered C_{H2} comprise the depreciated investment cost and all fixed and variable operational expenses for the entire supply chain here divided in export, PtX shipping, import and the CO_2 cycle in Eq. (9):

$$C_{H2} = \frac{c_{export} + c_{ship} + c_{import} + c_{CO2}}{m_{H2,supply}}$$

$$c_i = AF_i * CAPEX_i + OPEX_i$$

$$AF_i = \frac{(1 + WACC)^n * WACC}{((1 + WACC)^n - 1)} \quad (9)$$

with

- $\boldsymbol{CAPEX_i}$ = total capital expenses of all facilities based on their specific capital cost assumptions and capacities.
- $\boldsymbol{OPEX_i}$ = annual operational expenses (e.g. electricity, water, operation & maintenance, insurance).
- $\boldsymbol{AF_i}$ = specific annuity factor for capital expenses with the interest rate ($\boldsymbol{WACC}$) as weighted average cost of capital for the depreciation period $\boldsymbol{n}$ (usually applying a WACC of 8 % for a depreciation period of 20 years as stated in the ESI—S3).

In addition, the levelized cost of hydrogen (LCOH) (Eq. (10)) relate to the hydrogen produced at the electrolysis site. Thus, the LCOH replace only a share of the total expenses of the electrolysis plant in relation to the amount of hydrogen delivered considering energy losses and the PtX consumption as a ship fuel along the supply chain. The electricity consumption in Germany (e.g. pipeline compressor stations, reconversion plants) considers an electricity price of 50 € per MWh.

$$LCOH = \frac{AF * CAPEX_{EL} + OPEX_{EL}}{m_{H2,produced}} \quad (10)$$

3.9. GHG emissions intensity of PtX imports

In this study, the baseline assessment of the GHG emissions intensity of hydrogen supplied to an industrial site includes GHG emissions caused by the ship transport of the hydrogen carrier using either a conventional propulsion for fossil marine fuels (HFO) or engines that consume the PtX product as a fuel, the ship transport of CO_2 with dedicated vessels using HFO and the carbon capture process at the dehydrogenation plant according to Eq. (11) within a well-to-point of delivery system boundary as show in Fig. 1. The use of fossil marine fuels for shipping includes life-cycle emissions allocated to the supply of the fuel based on upstream emissions according to the GHG calculation methodology for RFNBO. [21] The emissions intensity levels are then compared against the maximum GHG emission intensity threshold of 3.38 kg CO_2(e) per kg H_2 for RFNBO. [21]

$$E_{H2} = \frac{(e_{td} + e_{u,fuel} + e_{p,release} + e_{leakage})}{m_{H2,supply}} \quad (11)$$

with.

- $\boldsymbol{e_{td}} = \boldsymbol{e_{td,WtT}} + \boldsymbol{e_{td,TtW}}$ = well-to-tank and tank-to-wake emissions for marine fuels used for the PtX and CO_2 ship transport based on Regulation (EU) 2023/1805 [115].
- $\boldsymbol{e_{u,fuel}} = \boldsymbol{e_{td,TtW}}$ = emissions from PtX fuel combustion during PtX ship transport.
- $\boldsymbol{e_{p,release}}$ = emissions from the PtX consumption for covering the energy demand of the PtX reconversion.
- $\boldsymbol{e_{leakage}}$ = emissions from hydrogen, PtX and CO_2 leakages (excluded in the baseline scenario).

All other process steps of the supply chain are excluded from the emissions accounting with the assumption of consuming exclusively renewable electricity and that the heat demand for dehydrogenation is covered by the imported PtX product without external fuels. In this context, it should be noted that additional GHG emissions can particularly arise from methane and CO_2 leakages in the eNG/CO_2 cycle including BOG losses during transport and storage, since methane is a strong greenhouse gas with a GWP100 of 25 g CO_2(e) per g CH_4. The impact of methane leakages is assessed in a leakage scenario with losses during the liquefaction and regasification process based on leakage measurements from Innocenti et al. with 0.070 % and 0.018 % of throughput at large-scale facilities. [116] It is assumed that CO_2 liquefaction and regasification account to similar leakage rates in this scenario. In addition, any hydrogen leakage potentially occurring during compression and storage or pipeline transmission is excluded in the base scenario applied for the cost assessment and included in an extended leakage scenario here. Leakage rates are considered with 0.5 % for hydrogen compression for the buffer tank storage and 0.5 % as an maximum estimate for pipeline transmission. [61] Note, that hydrogen leakage contributes to global warming with a reported GWP100 of 11.6 ± 2.8 according to Sand et al., while data on real-world leakage rates is scarce. [117–119]

The following assumptions were made for shipping. BOG losses during CO_2 shipping and storage are not considered in the baseline scenario. Any pilot diesel fuel use (e.g. HFO) in dual-fuel engines running on methanol, ammonia or e-NG is neglected here. [104] BOG generation for CH_4 shipping with a maximum daily rate of 0.07 % of the stored cargo for modern LNG carriers is captured and used as a fuel. [120] Direct methane emissions due to uncombusted methane in the engine exhaust gas, referred to as methane slip, are neglected. On this topic, Balcombe et al. present first direct measurements of GHG emissions from modern LNG carriers showing that methane contributes 35 % of the GHG emissions with an average engine slip rate of 3.8 % and specific total emissions of 10.11 g CO_2(e) per (t_{LNG} km) per unit of LNG

delivered for a one-way distance of 10,284 km (100-year time-horizon global warming potential (GWP100) of 36 g CO_2(e) per g CH_4). [120] This value is used as a benchmark in the results. Proposed onboard carbon capture systems for LNG-fuelled engines are not considered. [121] The LOHC(BT) carrier is alternatively assessed with a LNG-powered propulsion. GHG emissions of future ammonia engines only depend on the use of a pilot diesel fuel for ignition and combustion stability. Even so, the potential of GHG emission reduction depends significantly on nitrous oxide (N_2O) emissions of future engines and can be a barrier to adoption even with small amounts emitted, since N_2O is a strong GHG with a GWP100 of 265. [103] The European Maritime Safety Agency does not refer to specific emission values in their recent ammonia report. [103] A working paper from the Mærsk Mc-Kinney Møller Center for Zero Carbon Shipping estimates the impact of potential N_2O emissions with 0.06 g N_2O per kWh as a target emission threshold value for future ammonia engine designs corresponding to an emission factor of 0.158 g N_2O per kg NH_3 resulting into an GHG emission factor of 41.9 kg CO_2(e) per t NH_3, which is applied here. [122] All marine fuel emission factors and calculation formulae are given in the ESI (ESI-1).

4. Results

Since both scenarios are based on an equal electrolysis plant operation producing 90,429 t H_2 per year (3.01 TWh_{LHV}), the following results reveal the economic and environmental impact of different country-specific electricity costs and shipping distances on the final hydrogen supply volumes.

4.1. Economic assessment of hydrogen import scenarios

Australia – Germany. The energy equivalents for the specific PtX export and import volumes given in Fig. 2 reveal that the CH_4 route is the most efficient PtX-fuelled shipping mode consuming only 9 % of the exported PtX cargo, as it requires only 3 roundtrips with one vessel compared to 10 and 12 roundtrips with 3 vessels for ammonia and methanol, respectively. The higher synthesis losses of CH_4 are therefore compensated in a long-distance scenario with a slightly lower energy loss of 25.6 $\%_{LHV}$ in relation to the hydrogen production compared to ammonia (27.5 $\%_{LHV}$) and methanol (28.2 $\%_{LHV}$). After the dehydrogenation process at the import terminal, ammonia, LOHC and methanol yield though significant higher hydrogen supply efficiencies with 65.9 $\%_{LHV}$, 65.0 $\%_{LHV}$ and 62.7 $\%_{LHV}$ compared to CH_4 with 57.9 $\%_{LHV}$, since around twice as much energy is consumed for the reconversion of CH_4 compared to ammonia or methanol.

Consequently, the higher synthesis and reconversion efficiencies for ammonia and methanol lead to a much broader spread in the HFO scenario with a hydrogen supply efficiency of 78.7 $\%_{LHV}$ and 75.9 $\%_{LHV}$ for ammonia and methanol followed by LOHC(BT) with 65.0 $\%_{LHV}$ compared to CH_4 (63.1 $\%_{LHV}$). When assuming a lower ammonia cracking conversion efficiency (85.6 $\%_{LHV}$) as a main process uncertainty, the supply efficiencies of the NH_3 route are reduced to 62.1 $\%_{LHV}$ (PtX-fuelled) and 74.1 $\%_{LHV}$ (HFO). Detailed information about the energy quantities, respective plant and storage capacities, ship fuel consumption and the CO_2 export volumes are given in the ESI (ESI-4).

Incorporating the energy efficiencies into the economic assessment with the total capital and operational expenses results in the hydrogen supply cost visualized for both ship fuel modes in Fig. 3. Ammonia enables clearly the most cost-efficient hydrogen supply for both ship fuel modes, as the total costs range from 5.07 € per kg H_2 for ammonia (HFO) to 7.69 € per kg H_2 for LOHC(BT). The total costs exceed the LCOH of 2.24 € per kg H_2 in Australia by 226 % (NH_3) and 343 % (LOHC(BT)) in the lowest and highest cost scenario. The breakdown of the total electrolysis costs in Fig. 3 shows the shares of the LCOH and the additional electrolysis expenses allocated to the energy losses for synthesis, fuel consumption and dehydrogenation as depicted in Fig. 2. The total electrolysis costs range between 45 % for LOHC(BT) and 61 % for ammonia (PtX fuel) and are mainly made up of the electrolysis' electricity cost replacing between 27 % (LOHC(BT)) and 36 % (ammonia) of the total cost of the final hydrogen supply. Thus, a relatively low electricity price (25 € per MWh) remains still the main cost driver for all PtX pathways except for LOHC(BT) with a similar cost share for shipping.

The cost shares for the steady-state synthesis of the hydrogen carrier and the required hydrogen buffer storage are relatively low with down to 6 % for LOHC(BT) except for ammonia with up to 20 % due to higher capital cost and electricity expenses for the HB and ASU plant. The LOHC (BT) purchase cost for the for three operating vessels is thereby a significant cost driver replacing 83 % of the capital cost of the electrolysis plant. This shows a limited feasibility of LOHC(BT) for GW-scale hydrogen supply volumes over long distances. The CO_2 cycle is similarly a main cost driver affected by long-distance routes with up to 19 % and 24 % of the total cost for an annual export of 494 and 651 Mtpa CO_2

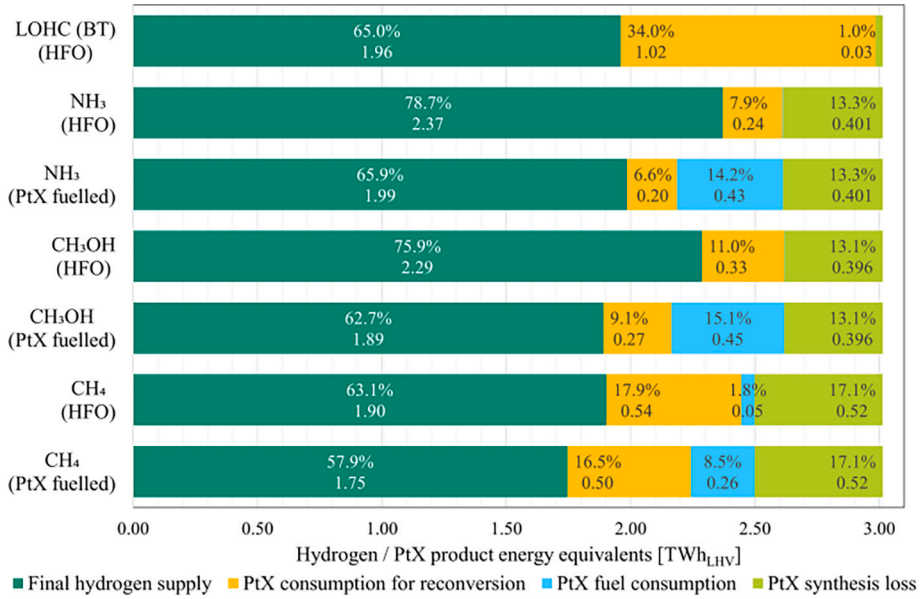


Fig. 2. Energy efficiencies of PtX products and final hydrogen supply in relation to the renewable hydrogen production in Australia.

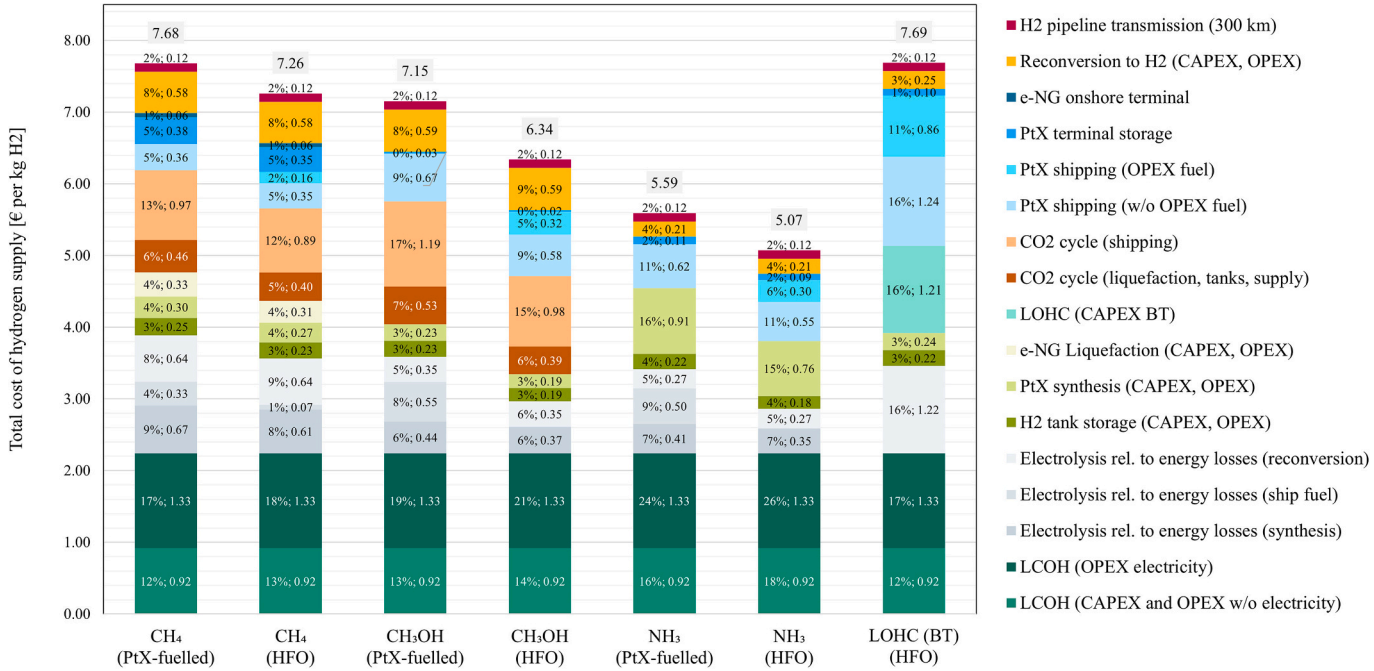


Fig. 3. Total hydrogen supply cost in the Australian scenario. The total electrolysis expenses consist of the shares of the levelized cost of hydrogen (green blocks) and the remaining electrolysis expenses (grey blocks), which are allocated to the PtX energy losses for synthesis, shipping and dehydrogenation. All other cost shares refer directly to the final hydrogen supply. All PtX routes except LOHC are differentiated by the propulsion mode of the PtX carrier consuming either the PtX product or heavy fuel oil as a marine fuel. (For interpretation of the references to colour in this figure legend, the reader is referred to the web version of this article.)

for CH_4 and methanol, respectively. This results into specific CO_2 supply cost of 151 and 150 € per t CO_2 for CH_4 and methanol (PtX-fuelled).

While the PtX shipping costs including the terminal storage are nearly similar for ammonia, methanol and CH_4 for both ship fuel modes, LOHC(BT) shipping is significantly more expensive due to higher fuel expenses (0.86 € per kg H_2) and charter rates (0.81 € per kg H_2) in particular. Shipping e-NG is in contrast the most cost-efficient pathway with the lowest PtX fuel consumption and without considering the terminal expenses. All fossil marine fuel scenarios reduce the total supply cost due to higher PtX import volumes outweighing the fuel cost of 0.16 (CH_4), 0.32 (methanol) and 0.30 € per kg H_2 (NH_3). A detailed breakdown of all shipping costs is given in the ESI (ESI-4).

Note, that often less noticed expenses like the Suez Canal fee contribute up to 0.24 € per kg H_2 for the methanol route with 0.12 € per kg H_2 each for the methanol and additional CO_2 carriers.

Dehydrogenation plant CAPEX replace a relatively small cost share below 9 % for all scenarios. Note, that the hydrogen losses are already priced in all other cost shares, which is highlighted by the breakdown of the electrolysis expenses in addition to the levelized cost of hydrogen. The electrolysis expenses allocated to dehydrogenation are highest for LOHC(BT) followed by CH_4 according to the conversion efficiencies. Considering a lower ammonia cracking efficiency, increases the total cost by 6 % to 5.91 (PtX-fuelled) and 5.36 per kg H_2 (HFO). The total cost of the imported PtX products disregarding the reconversion into hydrogen show a narrower cost range with 144 € per MWh_{LHV} for ammonia (741 € per t NH_3), 162 € per MWh_{LHV} for e-NG and 169 € per MWh_{LHV} for methanol (934 € per t CH_3OH) in the PtX fuel scenario (see ESI-4). This promotes a direct use of e-NG to avoid high energy losses related to the reconversion. This could tough increase the CO_2 cycle costs depending on the specific point of demand and end-use application under the assumption that a sufficient carbon capture rate is achieved (e.g. power plants with oxyfuel combustion or a direct-reduced iron plant equipped with CCUS). The pipeline transmission cost contributes 2 % of the final supply cost based on recent EHB assumptions, which highlights the cost efficiency of a highly utilized pipeline network in the medium term.

Tunisia – Germany. The effect of the PtX consumption for shipping is significantly lower for a short-distance route as shown in Fig. 5 as only 3.1 % and 2.9 % of the exported PtX cargo are consumed for methanol and ammonia. This leads to higher hydrogen supply efficiencies, especially for methanol (73.2 $\%_{LHV}$) and NH_3 (76.2 $\%_{LHV}$) compared to a long-distance route, whereas LOHC(BT) still surpasses CH_4 (63.2 $\%_{LHV}$). This correlates with reduced shipping expenses for the PtX import and CO_2 export, and leads to the largest relative cost reductions compared to Australia for methanol with 27 % (PtX fuelled) and 19 % (HFO), followed by LOHC(BT) (22 %) and CH_4 with 19 % (PtX-fuelled) and 15 % (HFO), while the ammonia route is less affected by the shipping distance with 13 % (PtX-fuelled) and 6 % (HFO) lower supply costs based on higher LCOH of 2.51 € per kg H_2 in Tunisia (see Fig. 4). Consequently, the cost range for the imported PtX products is relatively small with 117 € per MWh_{LHV} for methanol (655 € per t CH_3OH), 123 € per MWh_{LHV} for ammonia (637 € per t NH_3) and 128 € per MWh_{LHV} for methane (PtX-fuelled) (see ESI-3). Short distances clearly enhance the economic feasibility of the CO_2 transport cycle, which leads to a cost-efficient CO_2 supply with 49 and 54 € per t CO_2 for CH_4 and methanol in Tunisia. Note, that the LOHC(BT) route is clearly more applicable for short-distance supply chains, as the shipping expenses are 72 % lower for Tunisia compared to Australia. Moreover, the capital cost for the required LOHC material (BT) can be reduced by 50 % accordingly, since only one transport vessel is required and the BT bulk storage demand at the exporting site is reduced. This highlights a significant higher cost efficiency of LOHC(BT) supply. The impact of the different shipping fuel modes on the total cost can nearly be neglected with an increase of around 1 % compared to an increase of 13 % (methanol), 10 % (NH_3) and 6 % (CH_4) in the Australian scenario chains for short-distance routes and GW-scale hydrogen export volumes. The hydrogen consumption for LOHC dehydrogenation is then the main cost driver with 23 % of the electrolysis expenses after the levelized cost of hydrogen. This underlines the downstream cost reduction potential, in particular for LOHC (BT), when dehydrogenation plants could utilize waste heat that is potentially available from other industrial facilities at ports.

With respect to a lower ammonia cracking efficiency, the total cost

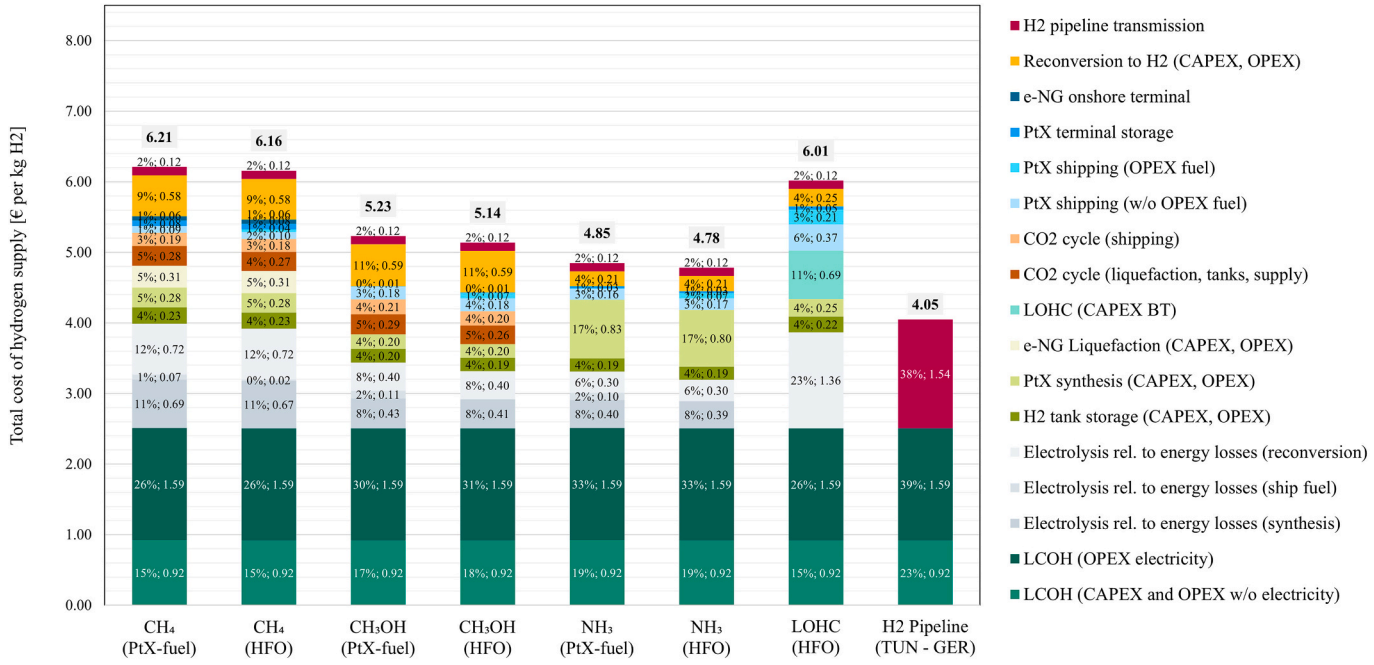


Fig. 4. Total hydrogen supply cost in the Tunisian scenario. The total electrolysis expenses consist of the shares of the levelized cost of hydrogen (green blocks) and the remaining electrolysis expenses (grey blocks), which are allocated to the PtX energy losses for synthesis, shipping and dehydrogenation. All other cost shares refer directly to the final hydrogen supply. All PtX routes except LOHC are differentiated by the propulsion mode of the PtX carrier consuming either the PtX product or heavy fuel oil as a marine fuel. (For interpretation of the references to colour in this figure legend, the reader is referred to the web version of this article.)

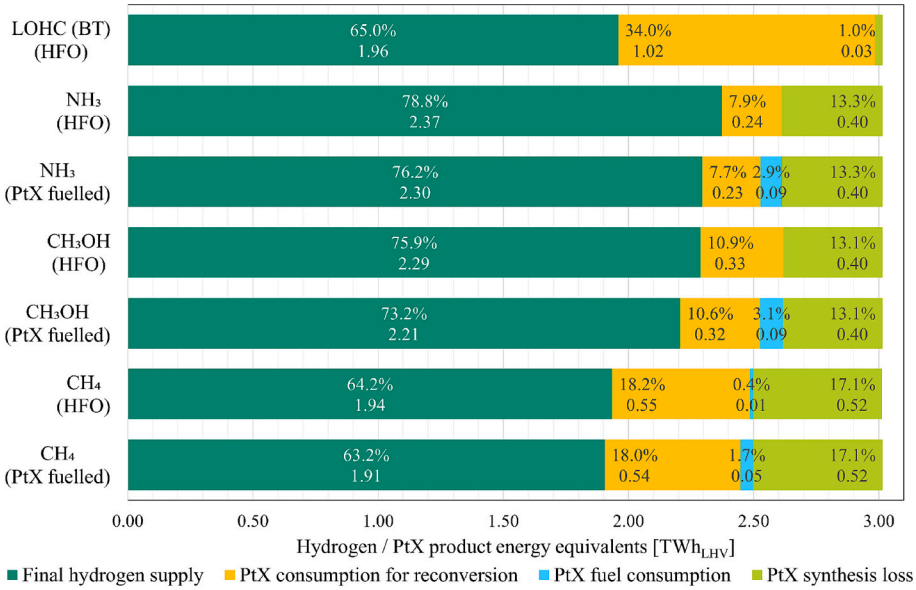


Fig. 5. Energy efficiencies of PtX products and final hydrogen supply of ship import routes in relation to the renewable hydrogen production in Tunisia.

increase by 6 % to 5.13 (PtX-fuelled) and 5.05 € per kg H_2 (HFO).

4.2. GHG emissions intensity assessment of hydrogen import scenarios

In the Australian scenario, LOHC(BT) and methanol clearly exceed the GHG emissions intensity threshold by 92 % and up to 44 %, respectively, as only the PtX-fuelled methane route and ammonia comply with the emissions threshold as shown in Fig. 6. Overall, the ammonia route causes the lowest GHG emissions even when running on fossil marine fuels before ammonia-fuelled engines become available. The LOHC(BT) route is environmentally not competitive for long-distance shipping routes, as the marine fuel consumption is 28 % higher compared to the CO_2/methanol cycle with additional CO_2 carriers. Assuming LNG-fuelled LOHC(BT) carriers reduces the total emissions by 24 %, which causes still a higher emissions intensity than for methanol. The LOHC(BT) route exceeds the emissions threshold at a distance of around 12,300 km or 16,100 km with Suezmax class tankers running on HFO or LNG on a well-to-wake basis, respectively. The CO_2 export is the main driver for increasing the emissions intensity with up to 55 % of the total emissions for the CO_2/methane cycle and high specific emissions of 2.29 kg CO_2(e) per kg H_2 for methanol (PtX-fuelled), which are nearly as high as for ammonia (HFO). This underlines the environmental benefit of deploying dedicated dual-use e-NG/CO_2 carriers. The CO_2/methanol cycle is thereby limited to a distance of around

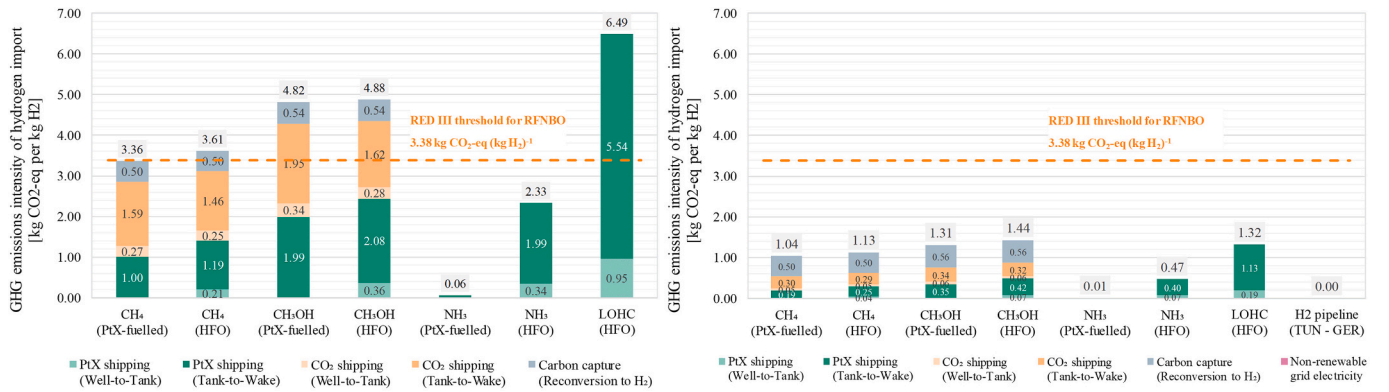


Fig. 6. GHG emissions intensities of hydrogen ship import routes to Germany (Port of Wilhelmshaven) considering different marine fuels for the PtX transport (left: Australia, right: Tunisia). Shipping routes: Tunisia, Port of Gabes, to Germany (4800 km), Australia, Port Kembla, to Germany (23,700 km). Direct pipeline route from Tunisia to Germany (4000 km) based on renewable electricity supply.

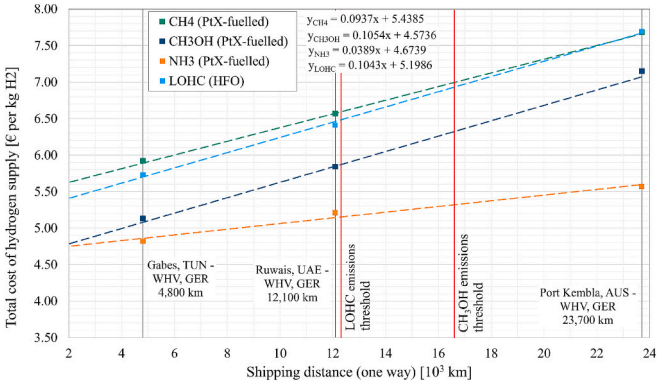


Fig. 7. Impact of shipping distance on total hydrogen supply cost under equal PtX production costs for a PtX-fuelled propulsion of PtX carriers. Dashed trend lines refer to average cost based on three shipping routes from Tunisia (TUN), Port of Gabes, United Arab Emirates (UAE), Port of Ruwais, and Australia (AUS), Port Kembla, to Germany (GER), Port of Wilhelmshaven (WHV).

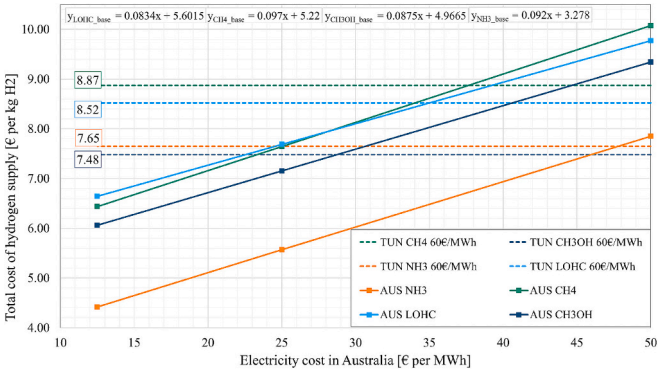


Fig. 8. Impact of electricity cost on the total hydrogen supply cost at an industrial site in Germany for a variable Australian scenario compared to a Tunisian high-cost scenario assuming electricity cost of 60 € per MWh (PtX-fuelled scenarios).

16,600 km. Achieving a high carbon capture rate of 94 % at the e-NG and methanol reforming plant effects a relatively low climate impact. The methane route profits from fewer required vessel roundtrips resulting in a 30 % lower emissions intensity compared to methanol in the Australian PtX-fuelled scenario. The e-NG transport accounts to specific GHG emissions per unit of imported e-NG of 13.7 and 12.5 g CO_2(e) per (t CH_4 km) for Australia and Tunisia (PtX-fuelled), respectively. These results can hereby be validated compared to real-world LNG carrier measurements of 10.1 g CO_2(e) per (t CH_4 km) based on efficient dual-fuel engines constantly consuming BOG. The climate impact of N_2O emissions for the ammonia-fuelled vessels is clearly negligible enabling nearly zero GHG emissions for the ship transport. The Tunisian scenario enables drastically reduced emission intensities for all carriers, so that the LOHC(BT) route and CO_2/methane cycle are well below the emissions threshold. In addition, it can be noted that the CO_2 cycle is in general no barrier to comply with emissions intensity thresholds. Assuming unabated CO_2 emissions from boil-off losses with a daily boil-off rate of 0.15 % restricts the methane and methanol route (PtX-fuelled) to distances of around 19,000 and 13,500 km, respectively.

Regarding the perspective direct hydrogen import route from Tunisia via pipeline to Germany an electricity consumption of 241 GWh is allocated to the compressor stations (ESI-4 Table 11), which needs to be exclusively covered by renewable electricity to account for zero GHG emissions. Assuming an electricity emission factor of 250 g CO_2(e) per kWh from non-renewable sources would result into a GHG emission intensity of 0.67 kg CO_2(e) per kg H_2 which is 43 % higher than the emission factor for the ammonia route with HFO (as shown in the ESI-3 Fig. 4). This underlines the need for integrating the sources of purchased electricity for long-distance pipeline routes into GHG calculation schemes. However, this accounts also for the downstream electricity consumption at the import terminal in Germany with a relatively high electricity consumption of 168 and 202 GWh for the methane and methanol route allocated to zero emissions in the GHG accounting.

The impact of hydrogen leakages and further GHG emissions arising from methane and CO_2 leakages is shown in an extended scenario in the ESI (ESI-3 Fig. 4), where the final hydrogen delivery is reduced by around 1 % for all ship routes, which increases the final supply cost accordingly. The climate impact of hydrogen leakage from pipeline transmission in Germany is thereby with 0.12 kg CO_2(e) per kg H_2 higher than the effect of potential methane leakages of the methane route during liquefaction and regasification with around 0.07 kg CO_2(e) per kg H_2. Total GHG emissions increase by 7 % for the Australian methane route exceeding the RFNBO threshold. The pipeline route from Tunisia to Germany accounts for total hydrogen losses of 2 % with an emission factor of 0.27 kg CO_2(e) per kg H_2 delivered to the offtaker, which increases the total supply cost given in Fig. 4 by 2 % accordingly.

4.3. Sensitivity of shipping distances on total supply cost

The impact of the shipping distance on the total hydrogen supply cost is shown in Fig. 7 assuming equal LCOH (2.24 € per kg H_2) for PtX-fuelled scenarios. Lower shipping expenses for shorter distances leverage only a minor cost advantage for LOHC(BT) compared to e-NG

due to the remaining high LOHC(BT) purchase cost for a hydrogen export of 3 TWh_{LHV}. The methanol route approaches thereby the ammonia cost range for short distances through reduced CO_2 transport cost. It is noteworthy, that the CO_2/methane and CO_2/methanol cycles leverage a cost-efficient CO_2 supply at the PtX site, especially for short and medium distances (95 and 101 € per t CO_2 for e-NG and methanol in the UAE scenario) compared to the future CO2 cost estimate of 200 € per t CO_2 from DAC. A prospective deployment of dedicated dual-use e-NG/CO_2 carriers could substantially increase the cost efficiency for e-NG by avoiding empty returns of the ships, while reducing the GHG emissions intensity especially for long distances.

4.4. Sensitivity of electricity cost on total supply cost

Since the electricity cost at the PtX production site remain the key cost driver, it can be stated that higher cost for longer shipping distances can be offset by lower electricity expenses as shown for a high-cost Tunisian scenario based on LCOH of 4.10 € kg per H_2 compared to an export from Australia in Fig. 8. All PtX routes successively achieve lower hydrogen supply cost below the electricity price limits of 47.5, 37.5, 35.0 and 28.7 € per MWh for ammonia, methane, LOHC(BT) and methanol, respectively. With decreasing cost to 12.50 € per MWh leveraging LCOH of 1.58 € per kg H_2 only the ammonia route undercuts the total cost in the Tunisian base scenario (in Fig. 8).

Additionally, reducing the electrolysis capacity to 1 GW has in contrast a limited total cost reduction effect with up to 5 % for ammonia, since the total electrolysis expenses are only slightly lower (6 %) than in the 1.5 GW base scenario due to stack replacement cost for higher FLH of 4800.

4.5. Perspective of direct hydrogen imports by pipeline compared to ship imports in the time period from 2030 to 2040

The final supply cost of 4.05 € per kg H_2 for a potential direct pipeline import route from Tunisia to Germany in an emerging pan-European hydrogen economy after 2030 are based on pipeline transport cost of 0.39 € per (kg*1000 km) for the recent EHB cost assumption (as shown in Fig. 4). This highlights the cost efficiency of direct hydrogen imports by pipeline, once potential cross-border import routes will become operational in the EHB over the next decade. A direct pipeline import yields 16 % lower final supply cost neglecting potential hydrogen transport losses compared to the ammonia ship route (PtX-fuelled), which questions the competitiveness of PtX supply chain by ship including long-term and capital-intensive investments in gigawatt-scale reconversion plants of ammonia or methane. It though remains uncertain, when and in what quantities hydrogen pipeline imports over long distances will be realized in the next decade, whereas maritime supply chains guarantee an early access to a diversified global export market in 2030.

4.6. Perspective of liquid hydrogen imports by ship and performance of this study compared to supply cost in literature

Recent studies estimate supply cost in 2030 of 6 € per kg H_2 for LH_2 compared to 7 € per kg H_2 for ammonia for a 500 tpd scenario (shipping distance 10,500 km) based on LCOH of 3.50 € per kg H_2 [40] and 5.28 € per kg H_2 for LH_2, while Rezaei et al. calculate an ammonia route with 4.32 € per kg H_2 based on LCOH of 2.40 € per kg H_2 (4 AUD per kg H_2) for a 1000 tpd scenario (shipping distance 7600 km) [36]. Hank et al. calculate even lower supply cost for LH_2 with 4.20 € per kg H_2 based on electricity cost of 25 € per MWh for the electrolysis plant (shipping distance 4000 km). [31] Note that a potential pipeline route between Tunisia and Germany shows 10 % lower cost with 3.79 € per kg H_2 when assuming electricity cost of 25 per MWh with the assumptions of this study compared to the lowest LH_2 scenario. The perspective deployment of large-scale LH_2 imports by ship remains thus highly uncertain from today's perspective. Reducing the specific energy consumption for liquefaction and scaling-up liquefier capacities (> 50 tpd) are key challenges to achieve liquefaction cost below 1 € per kg H_2 [18], while developing large-scale marine vessels. Note, that a LH_2 carrier would require a tank capacity of around 110,000 m^3 to store equal energy equivalents (~260 GWh_{LHV}) compared to a commercialized ammonia carrier with a 75,000 m^3 capacity.

The ammonia route in this study (PtX-fuelled) yields in comparison to the three selected studies and respective shipping distances (4000 km, 7600 km, 10,500 km) supply cost of 4.78, 4.97 and 5.12 € per kg H_2 based on LCOH of 2.24 € per kg H_2 (see Fig. 7). Compensating the different LCOH of the studies (2.40 and 2.25 € per kg H_2) shows an adjusted cost difference for the ammonia route (7600 km) of 0.81 € per kg H_2 higher cost compared to [36] (applying similar conversion and reconversion efficiencies with lower shipping expenses of 0.30 € per kg H_2 consuming ~24 % of exported ammonia, whereas the vessel in [40] is fuelled by a renewable marine fuel with Bio-LNG and SNG). Comparing the results for 10,500 km to [40] shows 0.62 € per kg H_2 lower cost (for ammonia as a fuel), while the ammonia routes in this study yields 14 % higher cost than LH_2 (4.20 € per kg H_2) in [31] for a distance of 4000 km. Major differences in parameter assumptions appear for instance for shipping in relation to the fuel and ship size (with capacities between 140,000–160,000 m^3 in [31,36]). Doubling the ammonia ship capacity in this study effects for instance around 5 % lower total supply cost, when assuming equal capital cost.

4.7. Limitations of the study design and results

The final results of the different hydrogen import options in this study are affected by the robustness of the techno-economic assumptions derived from different literature references with several uncertainties along the entire supply chain regarding the technology maturity, total cost and technical performance of gigawatt-scale PtX import projects. This refers especially to the respective conversion and reconversion plants for the hydrogen carriers, which are characterized by various uncertainties in terms of the capital cost of equipment, varying cost reduction estimates through economies of scale for future plant capacities and deviations between real-world and ideal energy conversion efficiencies reported in literature references. The assessment of PtX dehydrogenation plants for pure hydrogen supply is particularly derived from few literature references, such as the H21 industrial project report, and own calculations for covering the heat demand by burning hydrogen, which still needs further investigation and a detailed process simulation to validate the conversion efficiency for future plant designs. It is thereby important to integrate local factors, such as utilizing waste heat, to improve the conversion efficiency especially for LOHC(BT) with a relatively low energy efficiency compared to other hydrogen carriers, such as the applied ammonia cracking process. Additionally, the PtX import volumes and GHG emission intensities depend on the specific marine fuel consumption of the PtX ships in this study with varying engine power and corresponding ship speed, whereas the optimal tanker size for site-specific export volumes and optimal engine operation and fuel consumption (e.g. for methanol and ammonia engines) can differ from the assumptions in this study.

5. Discussion and conclusion

This study prioritizes the deployment of an ammonia-based GW-scale hydrogen import route from favourable green hydrogen production sites from overseas in terms of the final hydrogen supply cost and the transport-related GHG emissions intensity compared to LOHC(BT) and the proposed CCU-based CO_2/methane and CO_2/methanol cycles. These pathways are identified as the most relevant green hydrogen ship import options from overseas for short to medium term implementation, before liquid hydrogen could emerge as a feasible seaborne supply chain in the long-term. However, long-distance pipeline imports from low-cost

production countries can significantly reduce the cost of imports after an expected implementation in the 2030–2040 time-frame to meet a dedicated hydrogen demand in high-demand countries like Germany.

A limitation of the shipping distance is revealed for the LOHC(BT) and CO_2/e-methanol routes in terms of the GHG emission intensity especially underlining the need of defossilized marine fuels for an upscaled LOHC(BT) supply chain utilizing large crude oil tankers.

The main advantage of the CO_2/methane and CO_2/methanol cycles is the elimination of the shortage of feasible and sufficient climate-neutral CO_2 sources to deploy first GW-scale synthesis plants for carbon-based power-to-x products. All PtX routes are tough subject to various uncertainties associated to the technology-readiness-level of upstream and downstream assets and market forecasts whether a direct demand of PtX products restricts the reconversion into hydrogen to meet a dedicated demand of early industrial offtakers by fostering the emerging hydrogen pipeline network in Europe and Germany. The assumptions in this study address the following main techno-economic challenges and opportunities that should be a focus of future work. These include:

- A dynamic load operation of an integrated GW-scale PtX synthesis and electrolysis plant relying on an intermittent supply of renewable electricity to meet temporal correlation criteria for RFNBO based on site-specific data. This includes a hydrogen buffer storage system and a plant simulation with real-world PtX synthesis reactor efficiencies and effects on degradation and lifetime.
- The utilization of synergies between energy and heat flows through process intensification along the supply chains. On the upstream side, this is specifically linked to an integrated use of the excess heat from the synthesis reactor for additional power generation or high temperature electrolysis and direct-air-capture plants in the long-term. On the downstream side, this is linked to processes located at the import terminal (e.g. the integration of CO_2 liquefaction with the terminal regasification unit in the CO_2/e-NG cycle).
- The assessment of an industrial symbiosis concept between hydrogen import facilities and other industries located at import hubs. This is in particular aimed at utilizing waste heat for the heat demand of dehydrogenation (e.g. for LOHC or ammonia), which could be sourced from a gas-fired power plant or a direct-reduced iron plant as existing or new industries directly located at hydrogen import hubs).
- Reducing the techno-economic uncertainty of industrial scale ammonia cracking and methanol reforming plants equipped with CCUS and high capture rates (94 %) based on an internal heat supply consuming the PtX-hydrogen feed to fuel the dehydrogenation process.
- The deployment of dedicated dual-use e-NG/CO_2 carriers to increase the energy and cost efficiency of the e-NG/CO_2 cycle, while reducing the GHG emissions intensity.
- A detailed assessment of specific GHG emissions caused by the specific fuel consumption and potential BOG losses of the PtX and CO_2 carriers in comparison to real-world measurements of vessels.
- A market-oriented assessment of PtX value chains from an offtaker perspective comparing the specific willingness-to-pay for PtX products in existing markets (e.g. ammonia) with emerging new markets (e.g. hydrogen use in the steel sector) as a dynamic market tool and decision support for deploying new hydrogen import assets (e.g. hydrogen pipelines, underground storage or PtX reconversion plants).

At all events, implementing sustainable hydrogen import routes needs to comply with various Sustainable Development Goals beyond an internationally agreed accounting framework for GHG emissions along the supply chain. This includes environmental and socio-economic criteria and in particular a local value added, while fostering the energy transition and defossilization in the energy-exporting country during the deployment of a clean hydrogen export economy.

CRediT authorship contribution statement

Florian Scheffler: Writing – original draft, Software, Methodology, Investigation, Conceptualization. **Christoph Imdahl:** Writing – review & editing, Methodology, Conceptualization. **Sabrina Zellmer:** Writing – review & editing, Project administration, Funding acquisition. **Christoph Herrmann:** Writing – review & editing, Funding acquisition.

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

Additional data is given in the submitted Appendix.

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Appendix A. Supplementary data

Supplementary data to this article can be found online at https://doi.org/10.1016/j.apenergy.2024.125073.

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