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Strategies for large-scale deployment of low-emissions hydrogen for CO_2 abatement in petrochemical clusters

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Substantial amounts of low-emissions hydrogen are required to enable CO_2 emissions reduction and circularity in the petrochemical sector. However, large-scale deployment remains constrained by the persistent cost gap with fossil-derived hydrogen, the limited availability of low-cost renewable electricity and critical infrastructure, as well as the distinct operational constraints of hydrogen production technologies. This work addresses these barriers through technology diversification and integration of hydrogen production technologies. An integrated hydrogen production system is introduced that combines autothermal reforming with carbon capture and storage (ATR-CCS), solid oxide electrolysis cell (SOEC), and ammonia cracking (AC). A modeling framework is developed, centered on a mixed-integer linear programming model with a 1-year hourly resolution to optimize the technology mix and system operation while accounting for site-specific constraints and varying price conditions. The framework is demonstrated through a case study of a steam cracker plant, where integration provides opportunities to reduce hydrogen production costs via exports of displaced fuel gases. The results demonstrate how integrating multiple hydrogen production technologies reduces production costs while enhancing operational flexibility and system redundancy, compared with standalone systems. Scenario analysis highlights the importance of hedging against price uncertainty by deploying comparable or excess capacities across technologies to enable flexible operation. Based on these findings, a stepwise deployment strategy is proposed, outlining the timely implementation of individual technologies aligned with projected emissions allowance prices. The developed framework can be adapted to other clusters to identify cost-optimal, site-specific configurations for various market conditions, and thereby derive robust deployment strategies for scaling up low-emissions hydrogen production.

KEYWORDS
decarbonization, flexible operation, hydrogen, mixed-integer linear programming, optimization, petrochemical clusters

1 Introduction

Low-emissions hydrogen[1] is essential for achieving deep reductions in CO_2 emissions in sectors such as steel and ammonia production and is particularly important for enabling circularity in petrochemicals and transport fuels (IEA, 2023). It can be produced via electrolysis using renewable (green) or low-carbon grid electricity, as well as through steam methane reforming with carbon capture and storage (blue hydrogen). This hydrogen can replace both fossil-derived hydrogen and fuels used to meet industrial energy demands and is especially valuable at industrial sites where alternative CO_2 reduction methods, such as carbon capture and storage (CCS), are difficult to implement due to site-specific conditions or where direct electrification is unavailable or remains under development with uncertain timelines.

In this context, global hydrogen demand is projected to reach 150 Mt/year by 2030. In the European Union (EU), demand is expected to reach 45 Mt/year by 2050 (European Hydrogen Observatory, 2023), with near-term domestic production and import targets[2] (European Commission, 2025a). Despite this anticipated demand, large-scale deployment faces significant technical, economic, and systemic barriers (Odenweller and Ueckerdt, 2024). In particular, this deployment faces the so-called *chicken-and-egg* problem, where a coordinated scale-up of supply, demand, and supporting infrastructure is required (Schlund et al., 2022). However, in practice, this coordination is difficult due to a combination of cost-related and implementation prerequisites.

First, there is a persistent cost gap between low-emissions and grey hydrogen, primarily due to higher production costs and the low cost of emitting CO_2 (Odenweller and Ueckerdt, 2024). For example, although end-of-pipe CO_2 capture technologies are technically mature, their adoption in existing steam methane reformers (SMR-CCS) for blue hydrogen production remains limited. This reflects the continued viability of an *emit-and-pay* approach. CO_2 allowance prices as high as 100–200 €/tCO_2 translate to a cost penalty of just 20–40 €/MWh (~0.7–1.3 €/kgH_2). In addition, site-specific constraints, such as limited space and integration complexity, are key factors that increase costs and investment uncertainty for both retrofits and new installations (Roshan Kumar et al., 2025). Green hydrogen production cost estimates have also risen in recent years due to uncertainties in the future cost development of electrolyzers and limitations in their scalability (Ramboll, 2023). In addition, recent EU regulations require power purchase agreements (PPAs) for hydrogen to qualify as renewable, with certain conditions and exceptions (European Commission, 2023a), limiting cost optimization through flexible operation. Such optimization is possible only when procuring electricity on the spot market, albeit with greater uncertainty in electricity prices and associated emissions intensity.

Second, the availability of critical infrastructure determines when hydrogen production technologies can be deployed. For example, blue hydrogen production cannot be implemented until CO_2 transport and storage infrastructures are available. Similarly, the rollout of green hydrogen depends on access to reliable, low-cost renewable electricity, as well as critical infrastructure such as hydrogen networks and adequate grid transmission capacity. Moreover, the timing of such a rollout is crucial, as diverting renewable electricity to green hydrogen production before decarbonizing the power sector entails an opportunity cost (Mac Dowell et al., 2021).

Third, the risk of technology lock-in exists for both blue hydrogen and green hydrogen, although the degree of this risk depends on whether the hydrogen plant is tied to an isolated host plant or integrated with a plant located in a cluster with multiple alternative off-takers (Odenweller and Ueckerdt, 2024; Roshan Kumar et al., 2025). These deployment barriers indicate the need for a more comprehensive and practical evaluation of when, where, how, and which hydrogen production technologies can and should be deployed.

Several systems-level studies have analyzed the cost-optimal design and rollout of hydrogen infrastructure, considering a wide range of hydrogen production technologies (Ganter et al., 2024; Kountouris et al., 2024; Terlouw et al., 2022). These studies typically adopt a greenfield approach, relying on aggregated spatial, temporal, and technological representations, thereby overlooking site-specific constraints relevant to deployment at existing petrochemical sites. There is, therefore, a need for site-specific studies to improve the robustness of system-level analyses. Other studies have focused on individual hydrogen production technologies, particularly their techno-economic performances in different contexts (Biermann et al., 2022; Cooper et al., 2022; Devkota et al., 2024; Walter et al., 2023). However, these evaluations have often been conducted in isolation, based on differing assumptions and system boundaries, limiting comparability. Consequently, the question of technology selection for specific sites remains unresolved, as site-specific conditions may warrant a combination of technologies that can be operated flexibly and in coordination. A few studies have explored the complexity of integrating multiple process technologies into industrial clusters while simultaneously optimizing their installed capacities and their flexible operation under varying market conditions (Ibrahim and Al-Mohannadi, 2023; Tiggeloven et al., 2025; 2023). However, such studies either rely on hypothetical cluster configurations or assume full cooperation and the absence of competing interests among industries within clusters, limiting relevance to real-world conditions.

In summary, the existing literature on large-scale hydrogen deployment reveals several crucial gaps. First, most system-level analyses assume greenfield deployment, thereby neglecting both the integration benefits and the complexity of installation within existing sites. Second, the focus on a single technology fails to

1 The term 'low-emissions hydrogen' is used in this work, in line with IEA (International Energy Agency, 2025), to refer to both renewable (European Commission, 2023a) and low-carbon hydrogen (Official Journal of the European Union, 2025).

2 These targets are linked to hydrogen produced from renewable energy sources (excluding biomass) (European Commission, 2025a).

capture the potential synergies and operational flexibility offered by integrating multiple hydrogen production technologies. Third, previous studies that considered more than one technology were conducted with an aggregated spatial resolution, which overlooks site-specific limitations relevant to actual deployment. Finally, no study has yet examined how site-specific constraints, combined with flexible operation in response to variations in electricity, methane, and EU Emissions Trading System (EU ETS) emissions allowance prices, influence the technology mix (or system configuration) and the operation of such systems. These gaps in framing and modelling approaches imply a need for a generalized framework capable of identifying the optimal technology mix, capacity, and operation strategy of hydrogen production technologies, considering technology-specific constraints, site-specific factors, and energy price uncertainty.

This study develops such a framework, focusing on the potential to scale up low-emissions hydrogen production in petrochemical clusters by leveraging untapped synergies within the cluster to overcome the aforementioned deployment barriers. Given the varied operational characteristics and resource requirements of different hydrogen production technologies, integrating multiple technologies enables the system to adapt to evolving site conditions and reduce investment risk. To this end, this study proposes an *integrated hydrogen production system* combining solid oxide electrolysis cell (SOEC), ammonia cracking (AC), and autothermal reforming with carbon capture and storage (ATR-CCS), which offer distinct advantages that enhance operational flexibility and system redundancy (Section 2.3).

A framework methodology centered on a mixed-integer linear programming (MILP) model is applied to determine cost-optimal hydrogen supply configurations while accounting for site-specific constraints. The model also captures the dispatch strategies under varying price conditions. The framework is demonstrated using a case study of a steam cracker plant in a chemical cluster in Sweden, whose ethylene production capacity is representative of a majority of steam cracker plants in the EU (Petrochemicals Europe, 2025). Building upon our previous work (Roshan Kumar et al., 2025), which identified hydrogen firing as the most favorable decarbonization option for steam cracker plants, this study considers hydrogen firing as the primary end-use, representing a practical near-term solution for scaling up low-emissions hydrogen supply within the cluster (Section 2.2). The case study provides detailed insights into the optimal technology mix, hydrogen production costs, and potential deployment pathways for the integrated hydrogen production systems and compares these with standalone hydrogen production technologies under similar site and price conditions.

Scenario and sensitivity analyses are used to evaluate how key site-specific constraints (such as grid capacity, ammonia storage, and methane availability) and price conditions (including electricity, methane, ammonia, and EU ETS allowance prices) affect the optimal technology mix and hydrogen production costs. The study also investigates whether the flexible operation of the SOEC can reduce costs when operated in tandem with ATR-CCS and AC under varied energy market conditions. Finally, the study proposes a stepwise deployment strategy for scaling up low-emissions hydrogen within petrochemical clusters, leveraging short-term policy incentives.

The main contribution of this work lies in the development of a framework for identifying cost-optimal hydrogen supply and dispatch strategies, accounting for site-specific constraints, technology-specific limitations, and demand-side considerations within petrochemical clusters. Its novelty lies in determining the value of technology diversification and integration for overcoming deployment barriers facing low-emissions hydrogen, assessing trade-offs between flexible and steady operation, and deriving stepwise deployment strategies for such systems. The framework, therefore, enables a realistic assessment of deployment potential for low-emissions hydrogen production and serves as a basis for other clusters aiming for carbon neutrality through a shared hydrogen infrastructure.

The paper is structured as follows: Section 2 presents the integrated hydrogen production system. Section 3 describes the optimization framework. Section 4 outlines the case study. Section 5 presents the results, followed by a discussion on deployment strategies in Section 6, and Section 7 presents the conclusions.

2 Integrated hydrogen production system

Figure 1 shows the integrated hydrogen production system considered in this study. The hydrogen demand in the cracker furnaces is met through a combination of technologies, namely, methane reforming with CCS[3], solid oxide electrolysis cell (SOEC), ammonia cracking (AC), as well as hydrogen recovered from the byproduct fuel gas using gas separation technologies such as pressure swing adsorption (PSA). The figure illustrates the integration of these technologies through the shared steam network and indicates the possibility of exporting surplus hydrogen and recovered methane to the surrounding cluster. It also highlights the key site-specific constraints relevant to deployment, including the ammonia storage capacity, grid transmission limits, and availability of on-site fuel gas. This system configuration reflects the general concept of a hydrogen hub, which is defined as a regional network consisting of hydrogen producers, connective infrastructure including storage, and cross-sectoral end users (Clean Air Task Force, 2025). Since the scope of this work is limited to the production technologies and their integration within a chemical cluster for a specific end-use sector, the

3 The term "methane reforming w/CCS" is used here as a generalized descriptor of methane reforming pathways in the integrated system, encompassing conventional SMR units in existing refineries as well as emerging technologies such as eSMR or ATR-CCS when implemented as new installations. In this work, ATR-CCS is selected due to its higher CO_2 capture rates (>95%) relative to SMR-CCS (Johnson Matthey, 2025), and to enable a consistent comparison with other hydrogen production technologies (e.g., ammonia cracking and SOEC) that involve new capital investments.

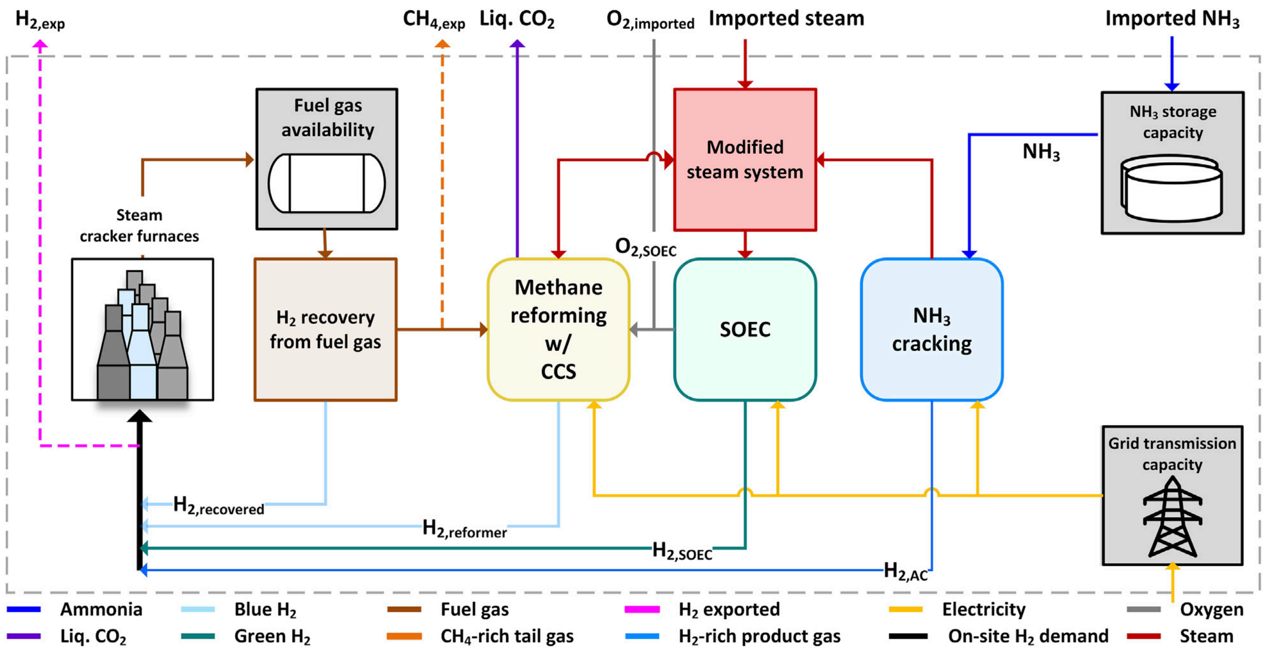

FIGURE 1
Simplified illustration of the integrated hydrogen production system, including three hydrogen production technologies: (1) methane reforming technology with CCS[3]; (2) solid oxide electrolysis cell (SOEC); and (3) ammonia cracking (AC). Hydrogen from these three technologies, together with the hydrogen recovered from the fuel gas, is used as fuel for the steam cracker furnaces. The possibility of exporting surplus hydrogen and recovered methane to the surrounding cluster is indicated with dashed lines. Gray-shaded boxes indicate the key site-specific constraints. For simplicity, the modified steam system is not shown here, although it is considered in the case study.

proposed system is hereinafter referred to as the *integrated hydrogen production system* or, simply, the *integrated system*. A more detailed description of this system is provided in Section 3.3.

2.1 Benefits of hydrogen deployment within petrochemical clusters

Petrochemical clusters are characterized by inherent redundancies that make them ideal sites for large-scale hydrogen deployment. First, the hydrogen demand in these clusters is substantial and time-invariant, whether for use as fuel or feedstock, which enables hydrogen to be supplied on a use-as-produced basis without the need for costly hydrogen storage systems. Consequently, hydrogen production technologies can achieve high capacity utilization or enhanced operational flexibility through overcapacity, as surplus hydrogen can be absorbed within the cluster. Second, these clusters provide access to multiple potential off-takers with existing interconnections for transporting industrial gases, steam, and other material streams. Specifically, the geographic proximity of industries within the cluster, along with existing fuel gas networks that could be repurposed for hydrogen transport, reduces the need for new hydrogen distribution pipelines. In addition, surplus steam from existing operations and newly installed processes can help to reduce the operating costs of high-temperature electrolyzers, such as the SOEC. Third, clusters typically have access to critical infrastructure, such as ports, natural gas grids, and power grids. Fourth, increased process electrification within the cluster would displace fuel gas combustion, enabling internal recovery of H_2 from the fuel gas and use of the resulting CH_4 as feedstock for blue hydrogen production or export to the natural gas grid. Fifth, industries within these clusters provide stable demand for low-emissions hydrogen in the short-to-medium term, but as conditions evolve toward a circular economy, this hydrogen can be redirected to other end-use sectors with higher CO_2 abatement potential or willingness to pay. Finally, a shared hydrogen infrastructure within the clusters helps minimize investment risks and supports phased deployment strategies.

2.2 Case study and rationale for hydrogen firing

Most of the CO_2 emissions from the steam cracker plant result from the combustion of fuel gas (methane and hydrogen) in the cracker furnaces and natural gas in the boilers. In our previous work (Roshan Kumar et al., 2025), post-combustion CO_2 capture was found to entail significant lock-in risks and to be particularly susceptible to cost escalation due to site-specific constraints. As an alternative, valorizing byproduct methane through methane reforming to produce hydrogen has been demonstrated as more cost-effective (Roshan Kumar et al., 2025), while also offering several practical advantages: i) hydrogen firing is relatively straightforward at this plant, as existing burners can already handle hydrogen-rich blends and, in some furnaces, pure hydrogen; ii) unlike post-combustion CO_2 capture, the integrated hydrogen production system does not risk becoming a stranded asset at the end of the lifetime of the steam cracker plant[4]. It also remains independent of future changes at the host site, such as the installation of e-crackers or recycling technologies; iii) the methane recovered from the displaced fuel gas, if injected into the natural gas grid, could qualify as *recycled carbon fuel* (European Commission, 2023a). Exporting this methane would directly displace natural gas imports in the EU and thereby reduce GHG emissions (European Commission, 2023b). This enables steam cracker plants to abate their Scope 1 emissions while simultaneously reducing the EU's dependence on imported natural gas. Additionally, the revenue from exported methane can help offset hydrogen production costs.

While hydrogen firing has clear advantages over post-combustion CO_2 capture, a more relevant comparison is with process electrification, widely regarded as the preferred long-term solution due to its substantial CO_2 abatement potential and the anticipated availability of cheap renewable electricity. However, large-scale demonstration of process electrification is still at an early stage, and commercial availability is not expected in the near term (Borealis, 2021). Moreover, retrofitting existing furnaces for direct electrification is expected to be relatively difficult compared to adapting burner systems for hydrogen firing. Finally, the availability of grid transmission capacity remains a key constraint for increased electrification, whereas hydrogen firing provides redundancy through fuel flexibility. In addition, unlike post-combustion CO_2 capture and electrified steam crackers, hydrogen firing is not linked to the residual lifetime of the host plant, allowing hydrogen production plants to be repurposed to supply low-emissions hydrogen to other off-takers. These advantages make hydrogen firing a more practical near-term solution for decarbonizing high-temperature heat supply in steam cracker plants. More broadly, the choice between direct electrification and low-emissions hydrogen to decarbonize high-temperature industrial heat (>400 °C) remains uncertain and is expected to vary across industries (Schreyer et al., 2024; Shafiee and Schrag, 2024). In this context, the use of green hydrogen for industrial heating remains contested due to competing end-uses and high abatement costs (>768 €/tCO_2) (Shafiee and Schrag, 2024; Terlouw et al., 2025), while blue hydrogen is expected to remain cost-competitive under favorable economic conditions and stringent operational standards (Ueckerdt et al., 2024).

4 The reason is twofold: on the demand side, hydrogen used as cracker fuel can be redirected to other industries in the cluster; on the supply side, if byproduct methane from the steam cracker plant is no longer available, natural gas can be sourced from the grid as a reformer feedstock.

2.3 Technology choices and limitations

Unlike blue hydrogen plants, which rely on external natural gas as the feedstock, an ATR co-located at a steam cracker plant would rely on methane recovered from fuel gas that was previously combusted as cracker fuel. As a result, repurposing this fuel gas as feedstock for the ATR does not introduce any additional operating costs. This eliminates a major cost component for blue hydrogen production, thereby motivating the inclusion of ATR-CCS in the integrated system. Although emerging technologies such as e-SMR (Wismann et al., 2019) can electrify the reforming process, they were not considered in this work due to their lower technological readiness levels and limited cost data. Nonetheless, deployment of either technology ultimately depends on access to CO_2 transport and storage infrastructure. Significant capital cost reductions can be achieved through economies of scale for such furnace-based hydrogen production technologies (Johnson Matthey, 2025). However, these technologies are designed for continuous operation and, therefore, have a limited operational range and slow ramping capabilities (Tiggeloven et al., 2025). In contrast, electrolyzer technologies, such as proton exchange membranes (PEMs), alkaline electrolyzers, and SOECs, offer operational flexibility by producing hydrogen when electricity prices are low and idling when prices are high. This motivated their inclusion in the integrated system.

SOECs were selected over commercially available low-temperature electrolyzer technologies based on several considerations. First, they offer high overall system efficiency (~75%, LHV basis), accounting for both electricity and steam supply. Second, unlike PEM and alkaline systems, SOECs can directly utilize steam as input, and this possibility enabled the use of surplus steam available on-site or the import of steam from existing or newly installed processes within the cluster. As a result, electricity demand per unit of hydrogen produced is reduced, yielding electrical efficiencies of up to 85% (LHV basis). Third, SOECs are reported to have the lowest standby load (>3%) among the three technologies, with cold start-up times of >10 h and a ramp-up from hot standby in approximately 10 min (Topsoe, 2024). Although their load-changing capability is slower than that of PEM and alkaline electrolyzers, which can respond within seconds, this hot ramp-up time was deemed practicable, given the availability of excess steam and high-temperature waste heat within petrochemical clusters (Clean Air Task Force, 2023). Fourth, SOECs are the most capital-intensive of the three electrolyzer technologies, with relatively shorter stack lifetimes and more frequent replacements (Clean Air Task Force, 2023). Therefore, SOECs, representing the higher-cost alternative, were selected to avoid underestimating electrolyzer costs in the integrated system. Additional considerations for electrolyzer deployment include the utilization of co-product oxygen, the limited potential for cost reductions through economies of scale (Ramboll, 2023), their competitiveness relative to direct process electrification, and the availability of renewable electricity and grid transmission capacity.

Ammonia cracking is included in the integrated system as a complementary pathway to supply low-emissions hydrogen using

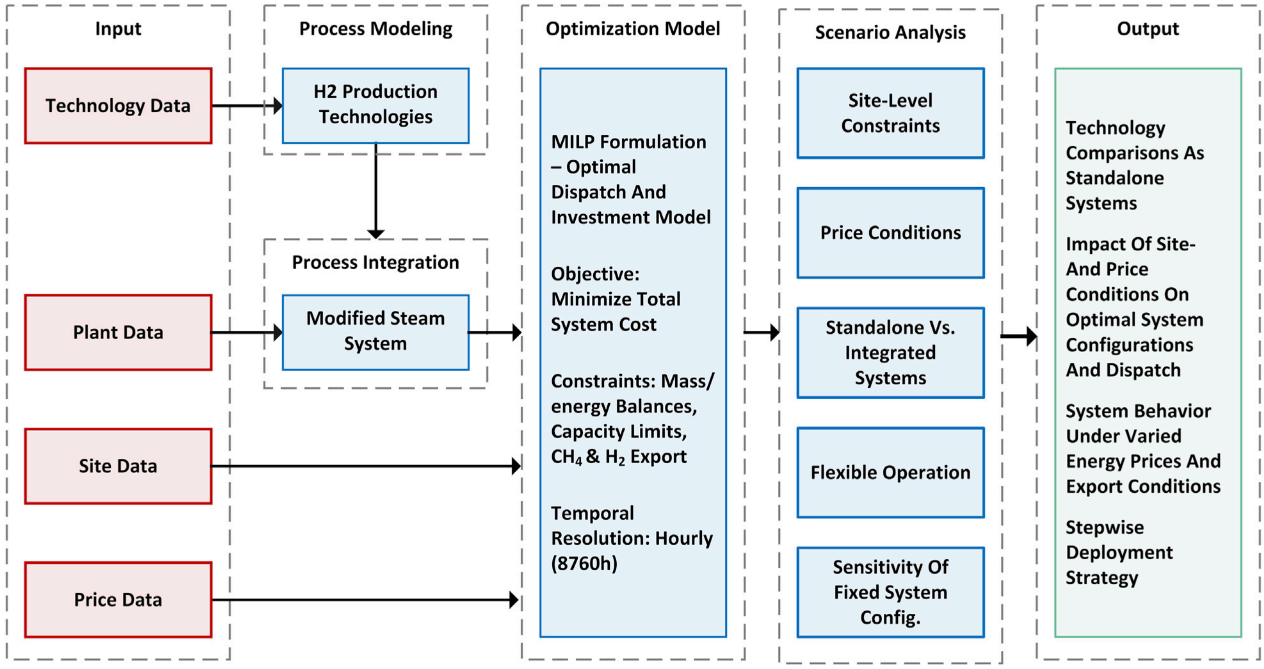

FIGURE 2
Overview of the framework methodology applied in this work. Red-shaded boxes indicate inputs, the green-shaded box indicates output, and blue-shaded boxes represent the numerical models developed and the scenarios investigated in this work.

either imported green ammonia from regions with abundant renewable energy or imported blue ammonia from regions where ammonia plants will be equipped with CCS. It offers several advantages: i) importing ammonia ensures security of supply and cost certainty through long-term contracts; ii) AC can be performed either on-site or on floating barges (Hoegh EVI, 2025), making it geographically flexible and a practical alternative for space-constrained sites; iii) in the case of blue ammonia, the use of cracked gas (mainly 75% H_2 and 25% N_2 by volume) as the primary fuel in the ammonia crackers decouples blue hydrogen production from the need for CO_2 transport and storage infrastructure; iv) AC offers fuel flexibility for the steam cracker plant, as ammonia can be used directly as fuel in the steam cracker furnaces. It may be blended with methane or fuel gas, partially cracked to form ammonia-hydrogen mixtures, or fully cracked to a hydrogen–nitrogen mixture before use as fuel in the furnaces (Mcallister and Henneke, 2024). However, using ammonia as a hydrogen carrier is fundamentally inefficient for end-uses such as electricity generation, renewable fuel production, or road transportation, yielding round-trip efficiencies below 20%[5] (Giddey et al., 2017). In contrast, in this context, when ammonia or its cracked gas is used as fuel in the furnaces, the corresponding round-trip efficiencies[6] are about 57% and 43%, respectively. Moreover, the capital costs for ammonia cracking represent only a small fraction (~10–12%) of the hydrogen supply cost, which is largely dominated by the ammonia feedstock cost (Lee et al., 2024). For this reason, AC is included in the integrated system as an alternative pathway to produce low-emissions hydrogen that complements ATR-CCS and SOEC by enhancing overall system flexibility and redundancy, particularly when site-specific constraints limit their deployment.

5 Including synthesis, transport, cracking, and the aforementioned end-uses (Giddey et al., 2017).

6 Assuming process efficiencies of 58.8% for ammonia synthesis, 75.9% for ammonia cracking including hydrogen separation, as well as 97% for transportation (Giddey et al., 2017).

3 Methods

Figure 2 illustrates the developed methodological framework to evaluate cost-optimal hydrogen supply configurations and their operation in petrochemical clusters. At the core of the framework is a cost-optimization model that enables assessment of technical interactions and economic performances across hydrogen production technologies subject to technology, site, and price constraints.

Steady-state process models of the hydrogen production technologies were developed to establish their mass and energy balances (Section 3.1). The hydrogen production technologies were integrated via a shared steam network based on their respective steam demands and potentials for steam generation through heat recovery (Section 3.2). The cost-optimization model was formulated as an investment and dispatch model to minimize the total system costs while remaining subject to a set of technical, operational, and site-specific constraints (Section 3.3). Key economic performance metrics were calculated from the optimization results to enable comparisons across scenarios (Section 3.4). These include the levelized cost of hydrogen (LCOH), which reflects the hydrogen production costs; on-site hydrogen supply cost (OHSC), which reflects the net cost of supplying hydrogen to a host plant, accounting for revenues from exports of surplus hydrogen and recovered methane; and the CO_2 abatement cost (CAC). The framework was applied to a reference steam cracker plant (Section 4–4.1). Scenario and sensitivity analyses were conducted to evaluate how site-specific constraints and economic conditions influence the optimal system configuration (Section 4.2), as well as how different energy pricing structures and hydrogen export conditions affect the operation of fixed system configurations (see Section 4 in Supplementary Material).

3.1 Process modeling

Detailed process models for the ammonia cracking process were adapted from previous work conducted by the authors (Axelsson et al., 2024), while ATR–CCS was based on (Roshan Kumar et al., 2025). Key technical and operational parameters for the SOEC were based on (Axelsson et al., 2024; Clean Air Task Force, 2023; Topsoe, 2024). The key process data and assumptions used for each hydrogen production technology are provided in Supplementary Table S1.

3.2 Process integration

The hydrogen production technologies were integrated via the steam system of the steam cracker plant, hereafter referred to as the modified steam system (see Supplementary Figure S2). The existing system was modified to utilize steam recovered from the ammonia cracker and ATR-CCS, as well as steam imported from the chemical cluster. A steam demand of approximately 9 tH_2O/tH_2 of medium-pressure (MP) steam (5 barg) was assumed for the SOEC subsystem (Topsoe, 2024). Process steam was not supplied directly to the SOEC stacks due to purity considerations. Instead, recovered MP steam was used to drive a dedicated steam generator supplied with demineralized water. Within the modified steam system, four potential sources of MP steam were available: i) the ATR consumed high-pressure (HP) steam and generated MP steam through heat recovery in the process gas coolers. Specific steam consumption and generation data were obtained from Roshan Kumar et al. (2025); ii) the cooling load profiles for the AC were used to estimate the specific MP steam generation potential (tH_2O/tH_2); iii) surplus low-pressure (LP) steam within the plant and imported steam from the cluster were upgraded using a mechanical vapor recompression (MVR) unit to MP steam conditions; iv) a standby electric boiler was included in the SOEC subsystem to enable operation during periods of MP steam deficit.

3.3 Optimization model

Figure 1 illustrates the major components, constraints, and system boundaries of the integrated hydrogen production system implemented in the optimization model. Here, methane recovered from the fuel gas is either used as feedstock for the ATR-CCS unit or exported to the natural gas grid. The resulting CO_2 stream is assumed to be liquefied to conditions suitable for ship transport (Northern Lights, 2025). A complete model description, including nomenclature, input data, assumptions, and all equations and constraints, is provided in the Supplementary Material.

3.3.1 Problem formulation

A multi-period MILP model with perfect foresight was developed to identify the optimal technology mix and hydrogen dispatch strategy for the integrated hydrogen production system. The model was implemented in the General Algebraic Modeling System (GAMS) and solved using the GAMS CPLEX solver with a branch-and-cut approach. The model takes a few minutes to solve on a standard laptop computer (Intel Core Ultra seven processor, 32 GB RAM) for simulating a full year of operation at hourly resolution. Binary variables were used to control the on/off status of a standby electric boiler. The objective was to minimize the total annual system costs while meeting the fixed hydrogen demand, subject to the technical and operational constraints associated with the modeled hydrogen production technologies. The model optimizes both investment and operational decision variables. The main constraints include the grid transmission capacity, ammonia tank capacity, fuel gas availability, minimum load limits, and ramp rate restrictions.

Objective function: Equation 1 defines the objective function of the model, which minimizes the total annual system cost of the integrated hydrogen production system.

$$\min C^{tot} = C_{inv,an} + C_{op,an} - C_{offset,an} \tag{1}$$

where c^{tot} is the total annual system cost, comprising the annualized investment costs ($C_{inv,an}$) the annual operational costs ($C_{op,an}$) and the total annual offsets ($C_{offset,an}$). Investment costs are annualized assuming a technology lifetime of 25 years and a discount rate of 10% for all technologies. These cost components are detailed in Equations 2–4, respectively. An annual maintenance cost of 5% of the annualized CAPEX was assumed. In addition, a 40% cost escalation was applied to reflect site-specific and retrofitting costs, in line with our previous study (Roshan Kumar et al., 2025).

$$C_{inv,an} = \sum_{n \in N} \left(C_n^{inv} \cdot i_n\right) + \left(C_n^{tank} \cdot \left(i^{tank} - i^{tank,existing}\right)\right) \tag{2}$$

where N is the set of all technologies, C_n^{inv} is the annualized investment of technology n (€/MW), and i_n is the installed capacity. The first term also includes investments in standby electric steam boilers and MVR, while the investment in additional ammonia storage capacity is included separately, which excludes the existing storage capacity ($i^{tank,existing}$). Equation 3 defines the total annual operational cost ($C_{op,an}$) as follows:

$$\begin{aligned} C_{op,an} = & \sum_{t \in T}\left(C_{AC,t}^{op} \cdot p_{AC,t} + C_{AC}^{cyc} \cdot \Delta_{AC}\right) \\ & + \sum_{t \in T}\left(C_{ATR\text{-}CCS,t}^{op} \cdot p_{ATR\text{-}CCS,t} + S_{ATR\text{-}CCS,t} \cdot c_{n,t}^{HPS} + C_{ATR\text{-}CCS,t}^{cyc} \cdot \Delta_{ATR\text{-}CCS,t}\right) \\ & + \sum_{t \in T} \sum_{n \in N_{steam}} \left(S_{n,t} \cdot C_{n,t}^{steam}\right) + \sum_{t \in T} e_t^{imp} \cdot \left(c_t^{el} + c^{tariff}\right) \\ & + \sum_{t \in T} q_t^{CO_2} \cdot c_t^{CO_{2,EUETS}} \end{aligned} \tag{3}$$

which includes technology-specific operational costs (C_n^{op}), cycling costs (C_n^{cyc}), steam costs, imported electricity (e_t^{imp}), as well as the CO_2

emissions costs e_t^{imp}. The technology-specific operational costs include feedstock costs, i.e., methane for ATR and ammonia for AC. For ATR-CCS, oxygen costs and CO_2 transport and storage costs for the captured CO_2 are also included. Imported electricity accounts for consumption in all relevant units, including the SOEC, compressors in ATR-CCS and AC, the MVR, and the electric boiler. CO_2 emissions costs are applied to residual emissions from ATR-CCS (~1%) and to the emission intensity of imported blue ammonia. The term $S_{ATR,t}$ denotes the HP steam demand in the ATR, while $S_{n,t}$ refers to the LP steam demands, primarily in the SOEC. Cycling costs are included in the optimization model to represent efficiency losses typically incurred due to load changes in process technologies. The term $\Delta_{n,t}$, denotes the positive load changes between timesteps for the ATR and AC, used to compute the cycling costs for these relatively slow technologies. The objective function also accounts for cost reductions from methane exports to the natural gas grid[7] and hydrogen exports to facilities within the cluster, as well as surplus upgraded steam that offsets natural gas use and, thereby, the CO_2 emissions on-site, as defined by Equation 4:

$$C_{\mathrm{offset}} = \sum_{t\in T} q_t^{\mathrm{CH_4,exp}} \cdot c_t^{\mathrm{CH_4}} + \sum_{t\in T} q_t^{\mathrm{H_2}} \cdot c_n^{\mathrm{H_2,exp}} + \sum_{t\in T} S_t^{\mathrm{surplus}} \cdot ei^{\mathrm{steam}} \cdot c_t^{\mathrm{CO_2,EUETS}} \tag{4}$$

The model formulation includes system-wide constraints for hydrogen, steam, electricity, and CO_2 balances, as well as technology-specific operational constraints (e.g., ramp rates, minimum loads, and ammonia storage dynamics). The hydrogen balance ensures that hourly hydrogen production meets on-site demand and export requirements, subject to installed capacity limits. Steam and electricity balances account for interactions between hydrogen production technologies and utility systems, including electric boilers and MVR. Methane balances track the allocation of available fuel gas between ATR-CCS demand and potential export. Net CO_2 emissions are calculated by accounting for both process emissions and avoided emissions from surplus steam utilization.

3.4 Economic evaluation

Based on the cost-optimal system design, two cost metrics were calculated: the levelized cost of hydrogen (LCOH) and the on-site hydrogen supply cost (OHSC). The LCOH is defined as the ratio of total annual system costs to total hydrogen produced (Equation 5). These costs include annualized capital investments and operational expenses, such as energy, maintenance, and emissions. The cost contributions of the electric steam boiler and the MVR were computed separately to allow comparisons of hydrogen production technologies across scenarios. The OHSC reflects the net cost of supplying hydrogen to the steam cracker plant and was calculated as the ratio of minimized total system costs (C_{tot}), including revenues from methane and hydrogen exports, to the total hydrogen produced (Equation 6). The corresponding CO_2 abatement costs (CAC) were calculated as the ratio of minimized total system costs to total CO_2 emissions avoided (Equation 7). Given the minor differences in absolute CO_2 abatement across technologies, the absolute CO_2 avoidance achieved with ATR-CCS ($\dot{m}_{CO_{2,avo}}$) was uniformly applied across technologies. Specifically, ATR-CCS was estimated to avoid approximately 610 $ktCO_2/y$ out of 651 $ktCO_2/y$ of on-site emissions. Detailed CO_2 avoidance calculations can be found in Roshan Kumar et al. (2025).

$$\mathrm{LCOH} = \frac{\mathrm{CAPEX_{annualized}} + \mathrm{OPEX_{yearly}}}{\sum_{t\in T}\left(p_{\mathrm{SOEC},t} + p_{\mathrm{NH_3},t} + p_{\mathrm{ATR-CCS},t}\right)} \left[\frac{€}{\mathrm{kgH_2}}\right] \tag{5}$$

$$\mathrm{OHSC} = \frac{\sum_{n\in N_{H_2}}\left(\mathrm{CAPEX_{annualized}} + \mathrm{OPEX_{yearly}}\right) - \sum_{t\in T} q_t^{\mathrm{CH_4,exp}} \cdot c_t^{\mathrm{CH_4}} - \sum_{t\in T} q_t^{\mathrm{H_2}} \cdot c_n^{\mathrm{H_2,exp}}}{\sum_{t\in T}\left(p_{\mathrm{SOEC},t} + p_{\mathrm{NH_3},t} + p_{\mathrm{ATR-CCS},t}\right)}$$

$$= \frac{c_{tot}}{\sum_{t\in T}\left(p_{\mathrm{SOEC}} + p_{\mathrm{NH_3}} + p_{\mathrm{ATR-CCS}}\right)} \left[\frac{€/y}{\mathrm{kgH_2}/y}\right] \tag{6}$$

$$\mathrm{CAC} = \frac{C_{tot}}{\dot{m}_{CO_{2,avo}}} \left[\frac{€/y}{\mathrm{tCO_{2,avoided}}/y}\right] \tag{7}$$

4 Case study–steam cracker plant

The steam cracker plant is located within a petrochemical cluster on the west coast of Sweden and processes ethane and other mixed hydrocarbon feedstocks. It has an annual production capacity of up to 640 kt ethylene and 200 kt propylene, representative of a majority of steam cracker plants in the EU (Petrochemicals Europe, 2025). Annual emissions are approximately 650 $ktCO_2$, with 80%–85% from cracker furnaces and the remainder from natural gas-fired steam boilers (12%–16%) and flaring (1%–3%). More details of the steam cracker plant and the surrounding cluster are provided elsewhere (Roshan Kumar et al., 2025; Thunman et al., 2019).

4.1 Assumptions for the case study

Hydrogen demand: The existing cracker furnaces operate with a fired duty of ~400 MW, corresponding to a hydrogen demand of 12 t/h (Roshan Kumar et al., 2025). After accounting for internally recoverable hydrogen from the fuel gas (~3.66 t/h), the net hydrogen demand was set to 8.34 t/h. Steam cracker plants typically operate continuously throughout the year, with annual maintenance shutdowns lasting up to 1 month and major turnarounds every 6 years (1–3 months) (Roshan Kumar et al., 2025). These shutdown periods were not represented in the model, and a constant hourly hydrogen demand was assumed over the year and across the technology lifetime.

Fuel gas availability: The case study plant currently produces approximately 500 MW of by-product fuel gas (50 vol% H_2 and CH_4), with 80% used internally as cracker fuel, while the remaining 20% is exported to downstream chemical industries (Roshan Kumar et al., 2025). Fuel gas availability can vary with changes in feedstock slate, downstream demand,

7 The tail gas from the PSA system for internal hydrogen recovery is primarily methane, with negligible amounts of light hydrocarbons (Roshan Kumar et al., 2025), and is therefore assumed to meet natural gas pipeline quality specifications.

electrification, or the decommissioning of existing crackers, as well as the integration of recycling technologies. To reflect this uncertainty, fuel gas availability was assumed to range between 250 and 500 MW, with the upper bound reflecting current operational conditions. The fuel gas composition was also assumed to remain consistent with current operations.

Electricity supply and transmission capacity: The case study plant currently imports electricity from the grid and has no direct connection to a co-located renewable electricity producer. Thus, renewable electricity was assumed to be imported from the grid. Hourly electricity price profiles from the SE3 bidding zone in Sweden for Year 2023 were used (Supplementary Figure S1), representing a year with high price volatility (–60 to 332 €/MWh; average ~50 €/MWh). Historically, wholesale electricity prices in this region have typically averaged around 50 €/MWh, with only a few hundred hours per year exceeding this threshold. Future wholesale electricity price estimates suggest that the annual averages may remain at comparable levels, with relatively low volatility owing to the large capacity of reservoir hydropower and substantial future demand for hydrogen in industry (Öberg et al., 2024).

The existing transmission capacity of the grid at the plant was estimated to be 100 MW (Borealis, 2023), representing approximately one-tenth of the total estimated transmission capacity to the regional grid in the Stenungsund region (850 MW) (ACCEL, 2024). Considering the expected grid expansion in the region (Svenska Kraftnät, 2024), a transmission capacity range of 100–500 MW was assumed in the scenario analysis.

Ammonia supply: The case study plant has access to an existing ammonia storage facility with a capacity of up to 5 kt (26 GWh) of ammonia (Argus, 2024). In addition, the plant has access to a port jetty, which could accommodate ammonia barges with capacities of up to 7–82 kt (40–480 GWh) (Hoegh EVI, 2025). Given that the port jetty could accommodate large ammonia barges and that large ammonia terminals typically have a capacity of 60 kt (Argus, 2024). The upper bound of the ammonia tank capacity in the case study was set to 60 kt (~312 GWh). A monthly tank filling frequency was assumed, corresponding to a filling rate of 4% of the total tank capacity (tNH_3/h), consistent with the combined filling rate of two marine loading arms (Baltic Chemical Terminal, 2015).

Surplus steam supply: The availability of surplus steam is highly site-specific. It typically varies across steam cracker plants and can fluctuate from year to year at the same plant. The case study plant has access to surplus steam from the surrounding cluster, which is condensed in the steam dump condenser during periods of excess supply. As described in Section 3.3.1, this steam is either upgraded and used to drive the steam generator of the SOEC or offsets steam production from natural gas boilers. Steam condenser data from the case study plant for a specific year were used to represent the availability and variability of surplus steam on-site.

4.2 Scenario and sensitivity analyses

The scenario and sensitivity analyses are divided into three parts. The hydrogen production technologies were first compared as standalone sources of hydrogen for the cracker plant, with their cost structures, LCOH, and OHSC ranges established through a sensitivity analysis (Section 4.2.1). Next, the impact of both site- and price conditions on the optimal system configuration and associated system costs was evaluated (Section 4.2.2). The impact of different energy pricing structures and hydrogen export conditions on the system behavior was evaluated, based on predefined scenarios combining fixed and time-resolved electricity and methane prices. A detailed description of the scenario design is provided in Section 4 of the Supplementary Material.

4.2.1 Comparison of H_2 production technologies as standalone systems

The technology-specific cost structures and the influence of price conditions and methane export were evaluated using the economic parameters listed in Table 1. A min–max sensitivity analysis was performed by setting each parameter to its respective lower and upper bounds, defining two extreme price scenarios—*low-price* and *high-price conditions*. These scenarios were used to establish plausible cost ranges (LCOH and OHSC) for the considered technologies. Site-specific constraints related to grid transmission, ammonia storage, and fuel gas were not applied to ensure fair comparison. In the case of SOEC and AC, the methane export potential was set to 278 MW, corresponding to the amount displaced by the hydrogen produced (8.34 t/h). In contrast, for ATR-CCS, methane in the fuel gas was used as feedstock rather than fuel. Accordingly, the OHSC included revenue offsets from methane exports for SOEC and AC, while methane feedstock costs were not considered for ATR-CCS. ATR-CCS, however, incurred additional costs for CO_2 transport and storage, assumed to be around 100 €/tCO_2 (COWI, 2021; Detterfelt and Hellström, 2024) for both price conditions.

4.2.2 Cost-optimal system configuration under varied site- and price conditions

The installed capacities of H_2 production technologies within the integrated system were constrained in the model by three key site-specific factors: grid transmission capacity, ammonia tank capacity, and the availability of methane recovered from fuel gas. Their impact on the cost-optimal system configuration was evaluated by running the model across a range of site-specific constraints and price conditions. Table 2 summarizes these site-specific constraints, which were based on the assumptions described in Section 4.1. The price conditions were defined using the economic parameter ranges in Table 1. The same max–min approach described in Section 4.2.1 was applied to determine the minimum and maximum possible system costs and the corresponding OHSC for the optimized configurations.

The scenario analysis was conducted using a structured scenario matrix based on a full factorial design. The grid capacity and fuel gas availability were mapped along the *x*- and *z*-axes, while ammonia storage was evaluated at two discrete levels along the y-axis. In total, 72 site conditions were evaluated and subsequently interpolated for enhanced visualization of their impact on the system configuration and on-site hydrogen supply costs. These scenario results were compared with the cost ranges estimated for standalone systems (Section 5.3) to identify site and price conditions under which lower OHSC and CO_2 abatement costs are lower than those of standalone systems.

TABLE 1 Economic parameters and their minima and maxima used to establish low and high price conditions, respectively. Note that CO_2 transport and storage costs were assumed to be the same under both price conditions (~100 €/tCO_2).

Parameters	Unit	Price conditions		References/Comments
		Low	High	
Ammonia price	€/tNH_3	400	1000	The assumed ammonia price range encompasses the reported price for blue ammonia (420–564 $/Mt) and renewable ammonia (781–818 $/Mt) (S&P Global, 2024). On an energy basis, the range corresponds to ~77–192 €/MWh. Emissions intensities of blue and green ammonia were set to 2.5 tCO_2/tNH_3 and 0 tCO_2/tNH_3, respectively (Axelsson et al., 2024)
Electricity price	€/MWh	50	100	Assumed based on the projected break-even PPA prices for new renewable electricity producers (S&P Global, 2023)
Methane price	€/MWh	30	90	This value is used for methane fuel, feedstock, or export. This range is also used to estimate the fuel costs avoided in scenarios with surplus steam from the ammonia crackers and the upgraded MP steam using MVR.
EU ETS emissions allowance price	€/tCO_2	100	200	On an energy basis, the assumed EU ETS allowance price range corresponds to ~20–40 €/MWh (methane) or ~0.7–1.3 €/kgH_2 expected, and is expected between 2026 and 2042 (Homaio, 2024; Mantulet et al., 2023)

TABLE 2 Assumed site-specific constraints used in the scenario analysis.

Site-specific constraints	Unit	Site conditions		References/Comment
		Low	High	
Grid transmission capacity	MW	100	500	The lower bound represents the estimated grid transmission capacity, based on current electricity consumption and existing grid connections at the steam cracker plant (Borealis, 2023)
Ammonia tank capacity	GWh	26	312	The lower bound represents the available storage capacity of 5 kt in the cluster, while the upper bound corresponds to large ammonia terminals with capacities up to 60 kt (Argus, 2024). The tank filling rate was capped at 4% of the total tank capacity, corresponding to 200 tNH_3/h at the lower bound and 2,400 tNH_3/h at the upper bound
Fuel gas (recovered methane) availability	MW	250	500	The upper bound represents estimated fuel gas production, including both exports as well as internal use as cracker fuel. Assuming the fuel gas consists only of methane, this range corresponds to 18–36 tCH_4/h (Roshan Kumar et al., 2025)

5 Results

5.1 Comparison of hydrogen production technologies as standalone systems

Figure 3 illustrates the levelized cost of hydrogen (LCOH) and on-site hydrogen supply costs (OHSC) for hydrogen production technologies as standalone systems under low-price and high-price conditions (see Table 2). Cost contributions are shown by stacked bar segments, while cost offsets from exporting recovered methane are highlighted in yellow. The OHSC values are indicated by black markers (×). For ATR-CCS, these markers represent OHSC calculated without methane feedstock costs, as this cost is already incurred when methane is used as cracker fuel.

Overall, the lowest LCOH range was estimated for ATR-CCS (2.44–5.27 €/kgH_2), followed by SOEC (4.13–6.87 €/kgH_2) and AC (3.92–8.42 €/kgH_2), based on the assumptions outlined in Table 2. These values correspond to a CO_2 abatement cost range of 292–630 €/tCO_2 for ATR-CCS, 493–821 €/tCO_2 for SOEC, and 468–1005 €/tCO_2 for AC, as per Equation 7. SOEC was estimated to be significantly more capital-intensive, with CAPEX contributing approximately 25%–42% to the estimated LCOH. In contrast, AC and ATR-CCS, which rely on similar process technologies, had CAPEX contributions below 18%. The LCOH also included a common cost component of 0.24 €/kgH_2 for recovering hydrogen and methane from the fuel gas. Operational costs were the main cost drivers across all technologies: electricity for SOEC (56%–67%), ammonia feedstock for AC (80%–93%), and methane feedstock for ATR-CCS (44%–60%). In addition to the operational costs, site-specific factors also influenced the hydrogen production costs. For example, ATR-CCS incurred additional costs for oxygen, transportation, and final storage of captured CO_2. The cost of oxygen typically depends upon its purity requirements, quantity, and the transport method

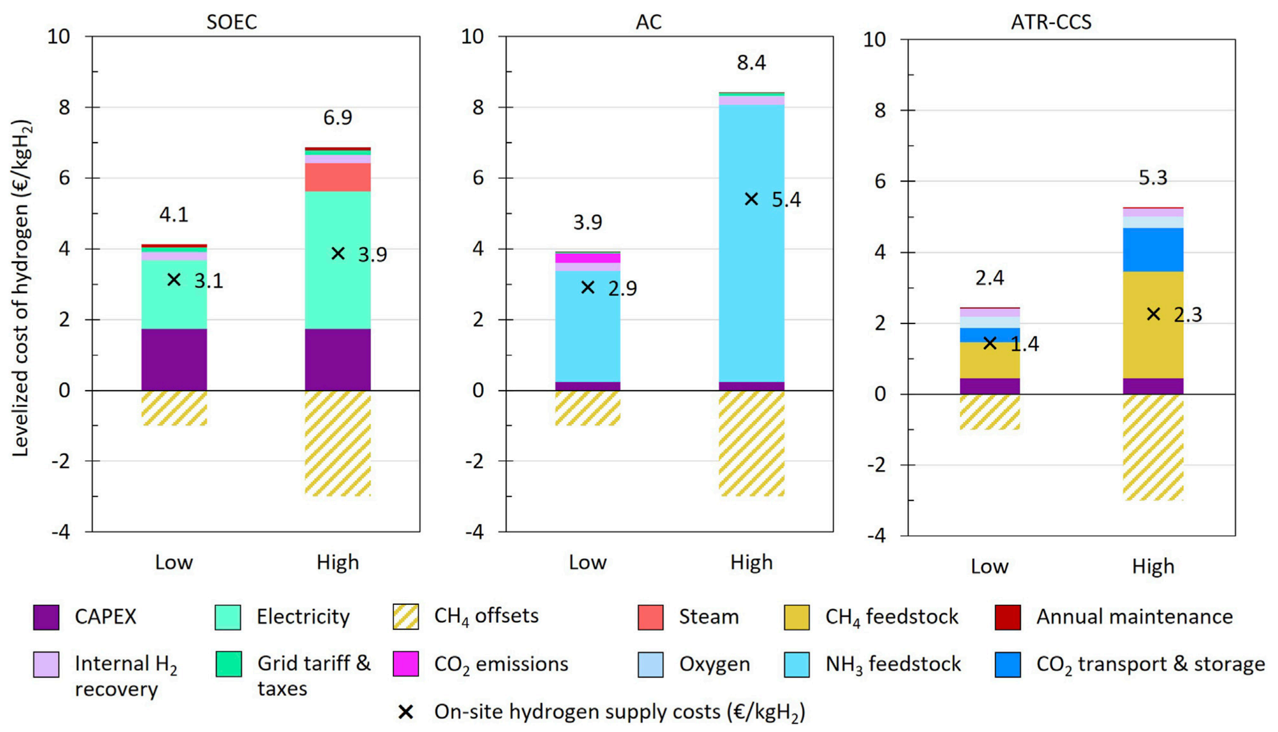

FIGURE 3
Levelized cost of hydrogen (LCOH) for standalone operation of solid oxide electrolysis cells (SOEC), ammonia cracking (AC), and autothermal reforming with carbon capture and storage (ATR-CCS) under low and high price. Diagonally hatched areas in yellow represent cost offsets from repurposing or exporting methane recovered from fuel gas. The solid black markers (x) represent the on-site hydrogen supply costs (OHSC), after accounting for these cost offsets.

(Gustavsson et al., 2023). Similarly, CO_2 transport and storage costs depend on the transport method used and the distance to the final storage site. In this case, proximity to an existing oxygen plant and an indicative CO_2 transport and storage cost of 104 €/tCO_2 (COWI, 2021; Detterfelt and Hellström, 2024) resulted in these cost components accounting for about 20%–30% of the total hydrogen production costs. Another relevant site-specific cost factor was the availability of surplus steam. As shown in Figure 3, steam costs could contribute up to 0.8 €/kgH_2 to the LCOH of the SOEC under high-price conditions, where the steam price reflects the upper-bound electricity price (100 €/MWh). This cost could be avoided entirely with access to excess steam from co-located industries in clusters.

Finally, the displacement of fuel gas and subsequent export of recovered methane enabled substantial cost reductions of up to 3 €/kgH_2 for SOEC and AC, corresponding to a methane price range of 30–90 €/MWh. A similar offset is possible for ATR-CCS if methane feedstock costs are not considered, as previously combusted fuel is now repurposed as feedstock for hydrogen production. With these offsets, the on-site hydrogen supply cost ranges for SOEC (3.13–3.87 €/kgH_2) and AC (2.92–5.42 €/kgH_2) narrowed substantially, making them comparable to that of ATR-CCS (1.44–4.03 €/kgH_2), which remained the least-cost option on average. These values correspond to CO_2 abatement cost ranges of 172–271 €/tCO_2 for ATR-CCS, 373–462 €/tCO_2 for SOEC, and 348–647 €/tCO_2 for AC.

5.2 Impacts of site and price conditions on integrated hydrogen production systems

Figure 4 illustrates the impacts of site-specific constraints, such as ammonia tank capacity, grid transmission capacity, and the availability of byproduct fuel gas, on the optimal system configuration, comprising SOEC, AC, and ATR-CCS, under low-price (Figure 4a) and high-price (Figure 4b) conditions.

The scenario analysis reveals two clear trends. First, access to more fuel gas increases the potential for methane export, significantly reducing OHSC. For instance, under low-price conditions (Figure 4a), the LCOH ranges between 3.0 and 3.7 €/kgH_2, while the OHSC ranges between 2.0 and 3.1 €/kgH_2. This offset is even more pronounced under high-price conditions (Figure 4b), where the LCOH is in the range of 5.6–7.9 €/kgH_2 and the OHSC from 1.2 to 5.7 €/kgH_2. Second, the optimal installed capacities were relatively consistent under low-price conditions, whereas under high-price conditions, significant changes to the system configuration were observed (see Supplementary Material). As a result, the increase in OHSC stems from site limitations that constrain technology choice, often leading to the deployment of hydrogen production technologies with higher costs.

Under low-price conditions, ATR-CCS dominates the optimal system configuration (Section 5.3, Supplementary Material). However, its installed capacity depends on the available ammonia storage capacity on-site, which enables the installation of larger

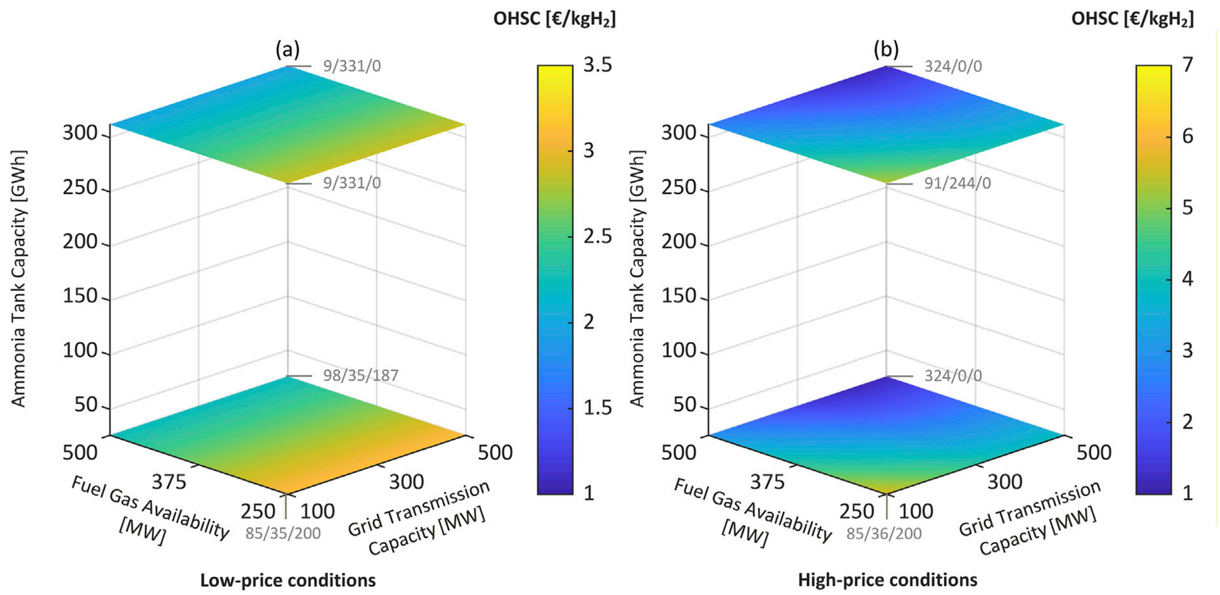

FIGURE 4
Impact of site-specific factors on the on-site hydrogen supply costs (€/kgH_2) under **(a)** low-price and **(b)** high-price conditions, accounting for offsets from repurposing or exporting methane recovered from the fuel gas. The optimal system configurations at the minimum and maximum estimated OHSC in the top and bottom planes of each subplot are annotated as SOEC/AC/ATR in MW, indicating the installed capacities of each technology (see Supplementary Material). Note: the OHSC scales differ between the two subplots. Additionally, these results are based on the economic parameters and assumptions listed in Table 1 of the Supplementary Material and are intended to illustrate trends in OHSC for site-dependent system configurations. These system configurations may change if alternate assumptions are applied.

ammonia crackers. When both ammonia storage and grid transmission capacity were limited, the integrated system shifted toward reformer-dominated configurations, regardless of energy price conditions. For instance, when ammonia storage is limited (bottom plane, Figure 4a), ATR-CCS supplied approximately 63% of the hydrogen demand, followed by SOEC (26%) and ammonia cracking (11%). This configuration, however, relied more heavily on internal methane use, which limits its ability to export methane. Conversely, when at least one of the two capacities (grid transmission or ammonia storage) was available, methane export was consistently favored, regardless of the methane price. When sufficient ammonia storage was available (312 GWh, top plane, Figure 4a), ammonia cracking became the primary hydrogen source, meeting 97% of the demand. The remaining share was supplied by the SOEC.

The shift from methane reforming to ammonia cracking as the dominant hydrogen supply option when there is sufficient ammonia storage can be attributed to two main factors. First, there is greater potential for reducing the total system cost through the export of byproduct methane in the integrated system configuration. Unlike in standalone ATR-CCS systems, where methane is repurposed as feedstock without incurring additional cost (Section 4.2.1), the optimized model treats methane use as a cost to ensure fair comparability. This reflects the opportunity cost of using methane internally rather than increasing the capacity of other technologies and exporting it. Second, ammonia cracking exhibits a slightly lower LCOH (3.68 €/kgH_2) than SOEC (3.88 €/kgH_2), as shown in Figure 3. This difference directly results from the price assumptions made for low-price conditions (Table 2). This also suggests that if renewable electricity could be procured under a PPA at a price <45 €/MWh, the optimal configuration would rely solely on the SOEC to meet the hydrogen demand, while simultaneously exporting methane. However, this outcome is only feasible when renewable electricity is available year-round, and sufficient transmission capacity is in place.

Under high-price conditions (Figure 4b), SOEC dominated the optimal system configuration when grid transmission capacity exceeded 260 MW. Below this threshold, AC dominated the configuration to both meet the hydrogen demand and minimize total system costs through methane exports. Although ammonia cracking exhibited the highest LCOH (8.4 €/kgH_2) among the three technologies under high-price conditions (Figure 3), this shift was primarily driven by the high methane price (90 €/MWh), which made exporting methane more economically favorable than using it as feedstock for ATR-CCS. Furthermore, when transmission capacity exceeded 420 MW, the optimal configuration consisted solely of SOEC, and nearly all methane was exported to minimize total system costs, which also corresponded to the lowest LCOH across all scenarios shown in Figure 4. However, this configuration is only optimal if by-product methane can be exported, as SOEC is not the least-cost option otherwise (Figure 3). The highest OHSC (>5.3 €/kgH_2) was observed in scenarios where both grid transmission capacity (100 MW) and fuel gas availability (250 MW) were limited.

5.3 Comparison of standalone and integrated hydrogen production systems

Figure 5 presents a comparison of the economic performances of the standalone hydrogen production technologies (Figure 3) and the

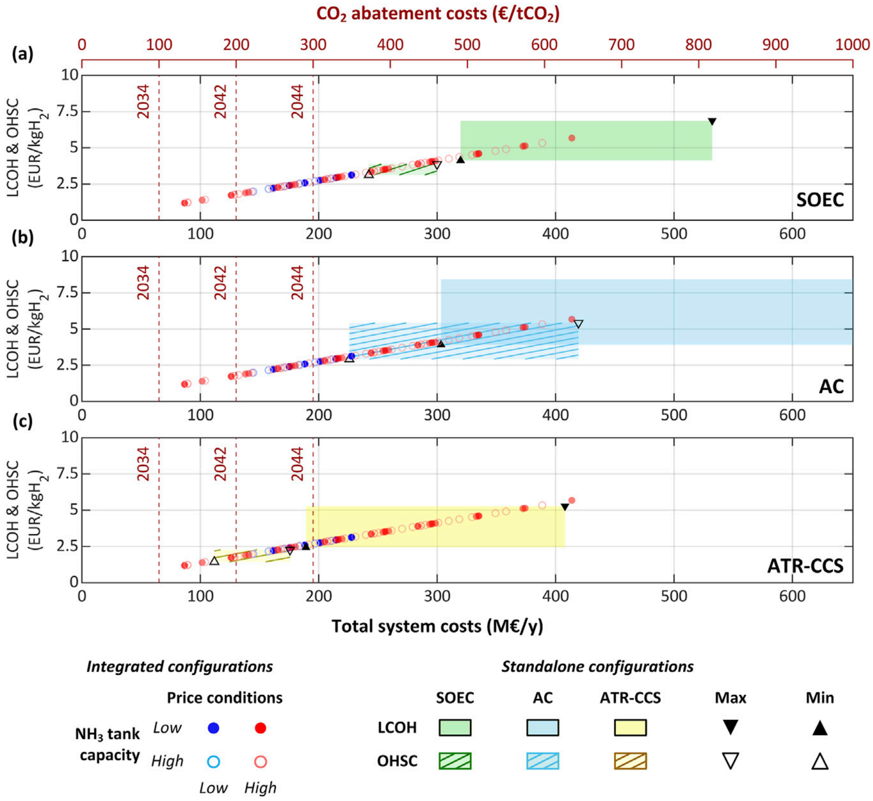

FIGURE 5
Comparisons of standalone hydrogen production technologies, **(a)** solid oxide electrolysis cells (SOEC), **(b)** ammonia cracking (AC), and **(c)** autothermal reforming with carbon capture and storage (ATR-CCS)—with optimal system configurations for different site and price conditions, and their relationships to total system costs, levelized cost of hydrogen (LCOH), on-site hydrogen supply costs (OHSC), and CO_2 abatement costs. Standalone systems are depicted as shaded regions, with diagonally hatched areas indicating costs accounting for offsets from repurposing or exporting methane. Optimal system configurations are represented by markers, where filled markers indicate limited ammonia storage (26 GWh), and unfilled markers indicate a larger storage capacity (312 GWh).

optimal system configurations obtained under varying site-specific constraints (Figure 4). For the standalone technologies, the estimated LCOH and OHSC ranges from Figure 3 are plotted against their corresponding total system costs, plotted as solid and hatched shaded regions, respectively. In Figure 5, the markers (○) indicate optimal system configurations (from Figure 4). Furthermore, Figure 5 illustrates the relationship between the total system costs and the corresponding CO_2 abatement costs (Equation 7). The vertical dashed lines indicate the model-based EU ETS allowance price projections reported by Mantulet et al. (2023).

Standalone configurations: Figure 5 shows that among standalone systems, ATR-CCS incurs the lowest total system costs (105 M€/year) and OHSC (1.44 €/kgH_2) under low-price conditions, corresponding to a CO_2 abatement cost of 172 €/tCO_2. In contrast, standalone SOEC and AC systems result in significantly higher abatement costs, ranging from 493 to 821 €/tCO_2 for SOEC and 468–1005 €/tCO_2 for AC. When accounting for cost offsets from methane exports, abatement costs remain prohibitively high, ranging from 373 to 462 €/tCO_2 for SOEC and 348–647 €/tCO_2 for AC. This implies that, as standalone systems, these technologies are economically unattractive, as they would likely require EU ETS allowance prices exceeding 300 €/tCO_2, not expected before 2044 based on modelled price projections (Mantulet et al., 2023). Notably, standalone ATR-CCS systems achieve comparable or lower abatement costs than post-combustion CO_2 capture (~243 €/tCO_2, including transport and storage costs) (Roshan Kumar et al., 2025). These costs are also comparable to those expected if the plant operates unabated and pays for emissions under the EU ETS (Mantulet et al., 2023). Together, these results indicate that conditions for deploying ATR-CCS as a standalone system could become favorable in the latter half of the 2030s, as increasing EU ETS allowance prices, driven by the scheduled phase-out of free allocations, are expected to render the *emit-and-pay* approach economically unviable.

Integrated configurations: In Figure 5, the degree of overlap between integrated system configurations (markers) and standalone systems (shaded regions) reflects their relative cost competitiveness. Three key observations can be made. First, the OHSC of integrated systems largely remains comparable to those of ATR-CCS, and to a lesser extent, AC and SOEC, as indicated by the overlap between markers and shaded regions. This indicates that, should site limitations (e.g., space or fuel gas availability) restrict the deployment of a standalone ATR-CCS, a site-dependent integrated system configuration becomes the next viable option. Second, a majority of the integrated system configurations lie below the lower bounds of the standalone system cost ranges. This is due to the broader sensitivity range assumed for fuel gas availability in the scenario analysis (250–500 MW), which exceeds the fixed methane export potential for standalone systems (~278 MW). As a result, more fuel gas can be exported under favorable transmission capacity conditions (>260 MW). Third, the availability of ammonia storage capacity has a negligible impact on overall costs, as indicated by the close positioning of filled and unfilled markers. This is linked to the limited deployment of AC across scenarios. Notable differences between the filled and unfilled markers emerge only when both fuel gas availability and transmission capacity are restricted, leading to AC-dominated configurations corresponding to the highest OHSC across scenarios.

Overall, the lowest costs are expected when favorable site and price conditions align. Favorable site conditions include (i) increased fuel gas availability, (ii) sufficient grid infrastructure to support large-scale SOEC deployment, and (iii) adequate ammonia storage to enable moderately sized AC when grid capacity is limited. Favorable price conditions include low electricity prices for SOEC, low ammonia feedstock prices for AC, and high methane prices that offset hydrogen production costs through methane exports. In practice, this implies securing price certainty through low-cost PPAs, long-term ammonia supply contracts, or attributing higher value to exported methane as recycled carbon fuels (see Section 6.1.4). Finally, given the uncertainty related to both site- and future price conditions, these results highlight the value of integrating multiple production technologies and aligning them with site-specific conditions to minimize system costs and reduce the risks associated with standalone systems.

Additional results on system operation under varying energy pricing and hydrogen export conditions, including dispatch and operational patterns, economic and CO_2 abatement performance, system configurations, capacity utilization, flexibility limitations, and the sensitivity of OHSC to key economic parameters, are provided in Section 5 of the Supplementary Material.

6 Discussion

6.1 Potential and implications of the integrated hydrogen production system

This work demonstrates the practical and economic advantages of integrating multiple hydrogen production technologies with steam cracker plants in chemical clusters over deploying standalone technologies. However, several limitations related to the scope and assumptions should be noted.

First, this study did not consider hydrogen leakage, which may affect both costs and the emissions intensity of low-emissions hydrogen. This assumption was based on the fact that the local fuel gas network in the chemical cluster routinely transports hydrogen-rich fuel gas (>50 vol% H_2) and typically exhibits low leakage rates due to regular monitoring and maintenance. However, leakage in broader hydrogen supply chains may be non-negligible (Alsulaiman, 2024; Esquivel-elizondo et al., 2023; Trapani et al., 2025). The estimated emissions intensities in this work are therefore indicative and depend on assumptions related to technology parameters (see Table 1; Supplementary Table S1). The use of established databases and methodologies is recommended for a more rigorous estimation of lifecycle GHG emissions (European Commission, 2023b).

Second, the practical deployment of the proposed integrated system at existing clusters depends on site-specific factors such as water availability and spatial constraints (ISPT and Deltalinqs, 2025; Manalal et al., 2025; Roshan Kumar et al., 2025). Furthermore, integration bottlenecks, such as compatibility of the existing fuel gas system with pure hydrogen, remain an important site-specific consideration. The costs associated with potential pipeline modifications or the installation of dedicated hydrogen pipelines to accommodate higher volumetric flows were not explicitly quantified. Nevertheless, these costs are expected to be minor relative to the estimated total system costs (Roshan Kumar et al., 2025) and represent a common cost component for both standalone and integrated configurations, and are therefore unlikely to affect the comparative cost estimates.

Third, the scope of the analysis was limited to the steam cracker plant, its hydrogen demand, and its utility balances. However, future process changes, such as the electrification of cracker furnaces, could enable repurposing low-emissions hydrogen as a feedstock for industries within the cluster. Such scope expansions must consider existing interconnections within the cluster to account for cascading impacts on material and energy flows (Tan et al., 2024).

Fourth, the modelling framework treats the integrated system as a price-taker, assuming its negligible influence on wholesale electricity prices. This behavior is expected in the short-term build-out phase (see Section 6.2). However, at larger deployment scales, aggregated electrolyzer loads could influence spot prices in the local electricity market, potentially transforming these facilities into price-setters (Kanellopoulos and Blanco, 2019). Additionally, constant efficiencies were assumed for all hydrogen production technologies, which do not capture partial-load behavior, potentially affecting the estimated cost results to some extent. This is particularly relevant for the SOEC (Schiedeck et al., 2025), which provides operational flexibility to the integrated system. Future adaptations of the framework could incorporate hydrogen leakage, spatial footprint requirements of new technologies, other site-specific constraints, cluster interconnections, partial-load performance of hydrogen production technologies, and the price impacts of large-scale electrolyzer deployment on local electricity markets.

The cost estimates in this study were based on a set of economic parameters to establish plausible cost ranges for each technology. Future technology cost projections were not considered to avoid underestimating capital costs. Operational costs, however, are likely optimistic due to the assumption of continuous year-round operation, which neglects maintenance shutdowns and variability in renewable electricity availability, particularly when directly connected to a renewable energy producer. As a result, the LCOH and OHSC ranges may be underestimated, particularly for SOEC, as reduced availability of renewable electricity lowers capacity utilization and increases production costs. In contrast to standalone systems, where low utilization is typically addressed through overcapacity and hydrogen storage, the integrated system avoids high upfront costs associated with storage by directly utilizing hydrogen as cracker fuel. Future work could enhance the framework by incorporating stochastic or robust optimization methods to identify robust system configurations for specific petrochemical sites. These analyses could also account for financial incentives for hydrogen production (European Commission, 2025b; US DOE, 2025) as well as electricity price projections across specific bidding zones (Energy Brainpool, 2024; ENTSO-E, 2022; Löfving et al., 2026).

6.1.1 Influence of site constraints on system configuration

The scenario analysis identified cost-optimal system configurations that minimize hydrogen supply costs under favorable site conditions. However, realizing such configurations may be challenging due to infrastructure and market-related uncertainties. Transmission capacity expansions typically require long lead times, while the availability of stable renewable electricity is not guaranteed, and sustained high prices for methane exports are unlikely without long-term contracts. Such limitations can be mitigated through more extensive deployment of AC and ATR-CCS within the integrated system, reducing dependence on grid availability. Configurations with relatively large reformer capacity restrict methane exports. In such systems, methane export would only be economically viable if a significantly higher value is attributed to the exported methane as recycled carbon fuel (see Section 6.1.4). Conversely, if greater operational flexibility with methane exports is desired, exposure to market price variations through relatively smaller reformer capacity[8] is required. Reformer sizing is therefore crucial, as it determines whether by-product methane is treated as an internal cost (feedstock for ATR) or as a potential revenue stream via exports. AC plays a complementary role at moderate capacities by enhancing flexibility, particularly when grid limitations constrain SOEC deployment. However, AC-dominated configurations increase the risk of technology lock-in, making system costs highly sensitive to ammonia price levels. These risks can be mitigated through long-term supply contracts.

6.1.2 The value of co-location, system integration, and overcapacity

Integrated configurations enable higher utilization of installed hydrogen production capacities by aligning operation with electricity price fluctuations and enabling strategic use or export of by-product methane to reduce both costs and emissions. Overcapacity in the integrated hydrogen production system is particularly valuable when export to the cluster is possible. In contrast to standalone electrolyzer systems, where flexible operation is made possible by coupling overcapacity with hydrogen storage infrastructure, the integrated system can rely on industries within the cluster to absorb surplus hydrogen production at an agreed-upon price. Assuming that the steam cracker plant, as the sole initial off-taker, invests in an integrated system with overcapacity, future exports to the cluster would increase capacity utilization and consequently reduce its OHSC over time. However, the extent and timing of the reduction in OHSC depend on when subsequent off-takers become available and the prices at which surplus hydrogen is sold. These considerations indicate that integrated systems with future off-take potential can help reduce investment risk while technology diversification hedges against changes in site conditions.

6.1.3 Trade-offs between price certainty and operational flexibility

Electricity and methane exhibit distinct trade-offs between price certainty and operational flexibility (Supplementary Figure S3). In the integrated system, operational flexibility is mainly enabled by SOEC, with ATR-CCS and AC playing complementary roles. Flexible operation, in response to electricity price fluctuations and surplus steam availability, can reduce system-wide costs. However, these outcomes will depend on how spot electricity prices evolve over time. In the EU, PPAs with renewable electricity producers are required to qualify hydrogen as fully renewable, while electricity sourced from the grid is subject to geographic and temporal correlation criteria (European Commission, 2023a). These requirements can lock in high prices over the long term and restrict operations to a specific bidding zone, thereby limiting operational flexibility and potentially reducing electrolyzer utilization. Future wholesale electricity prices are projected to remain volatile and may decline over time (Gasparella et al., 2023). Therefore, retaining operational flexibility may be more favorable than locking into fixed PPAs, which are typically priced well above 60 €/MWh (S&P Global, 2023). The operating strategy may also need to be adapted over time, considering evolving cost components such as grid transmission fees, which are expected to rise due to costly grid expansions (TenneT Netherlands, 2025). Furthermore, while renewable PPAs are mandated for producers of renewable fuels of non-biological origin (RFNBO), this requirement may not be necessary in this context, where low-emissions hydrogen is used solely as the cracker fuel. In such cases, the more relevant question is whether hydrogen must be fully renewable in the short term. In the Swedish context, with a grid carbon intensity of just 14 gCO_2/kWh (Bastos et al., 2024), SOEC could operate using grid electricity and still achieve the mandated 70% emissions reduction threshold. In the long term, however, classification as fully renewable may become necessary if hydrogen is later exported to other plants within the chemical cluster.

6.1.4 Methane export and its implications for CO_2 abatement

Exporting methane recovered from displaced fuel gas at the steam cracker plant offsets hydrogen production costs and reduces on-site CO_2 emissions. However, this raises an important question regarding the actual CO_2 abatement achieved, given that the exported methane remains fossil-derived. Nevertheless, the rationale behind this strategy is fourfold. First, the by-product methane exported from the steam cracker plant could displace fossil fuel gas imports to the EU, thereby contributing to energy security, as outlined in the REPowerEU strategy (European Commission, 2022). This implies that CO_2 emissions that would have otherwise occurred on-site are instead released elsewhere. For

8 Compensated by an increase in installed capacities of either AC or SOEC to meet the internal H_2 demand.

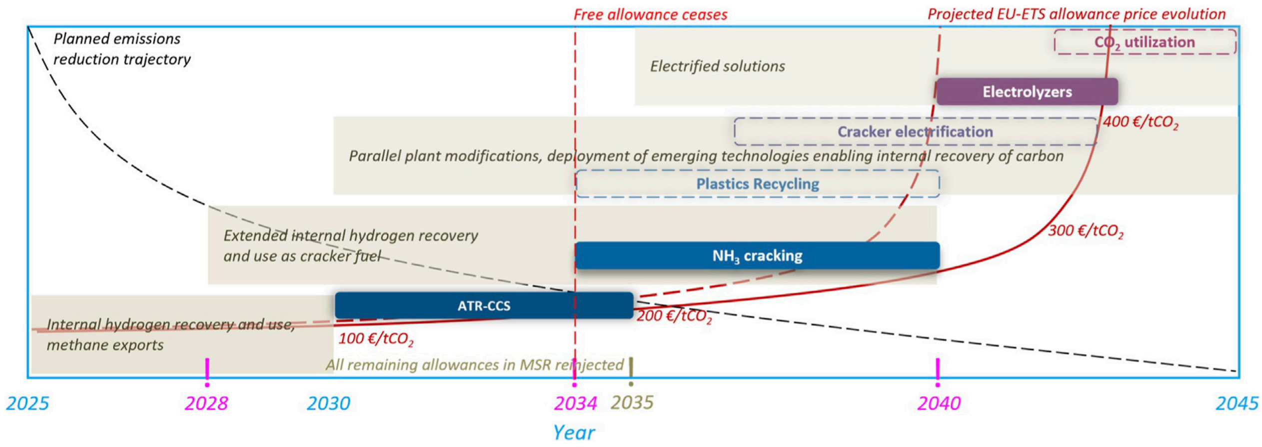

FIGURE 6
Indicative stepwise deployment strategy for a steam cracker plant within a petrochemical cluster, aligned with projected EU ETS price evolution (red curves), emissions reduction targets (black dashed curve), and scheduled turnaround years (pink markers). Two EU ETS allowance price trajectories are shown: one without reinjection of allowances from the Market Stability Reserve (MSR) (red dashed curve), and one with reinjection after the phase-out of free allowances in 2034 (red solid curve), based on (Mantulet et al., 2023). Technologies evaluated in this work are indicated with solid boxes, while technologies such as electrified crackers and plastics recycling, which could be deployed in parallel but are outside the scope of this work, are shown with dashed boxes. Note that this figure is illustrative and not drawn to scale.

the case study plant, such exports could offset up to 20% of Sweden's annual natural gas imports. When scaled to all steam cracker plants in the EU, under similar conditions, this corresponds to ~10% of the annual LNG imports and about 5% of total gas imports to the EU (European Commission, 2022) (Supplementary Table S6).

Second, GHG emissions estimates (Supplementary Figure S12) indicate that the exported methane exceeds the 70% threshold required for classification as a recycled carbon fuel (European Commission, 2023b). If the exported methane is considered as such, the associated CO_2 emissions are not attributed to the steam cracker plant, thereby strengthening the case for methane export as a valid short-term abatement strategy. This approach provides the necessary lead time for grid expansion and deployment of critical infrastructure, including CO_2 and hydrogen transport and storage supply chains. Third, the anticipated on-site transition supports maintaining flexibility between reforming for hydrogen production and export to the natural gas grid, both of which contribute to CO_2 abatement. Finally, this strategy supports the deployment of hydrogen production technologies in the short term, with the possibility to redirect hydrogen to alternative end-uses with higher CO_2 abatement potential within clusters.

6.2 Towards large-scale deployment of low-emissions hydrogen

A limitation of the proposed integrated system is that cost-effective implementation may require stepwise deployment of individual technologies, given that favorable site-specific and market conditions are expected to emerge over time. Practical implementation must therefore consider the order, deployment timing, and coordination within the cluster. The choice of technology ultimately depends on the cost of CO_2 abatement relative to the alternative cost of emitting CO_2.

Each technology becomes cost-effective at different EU ETS allowance price levels (see Figure 5): ATR-CCS (292–630 €/tCO_2), AC (468–1005 €/tCO_2), and SOEC (493–821 €/tCO_2). The overlapping abatement cost ranges suggest that these technologies compete within certain EU ETS allowance price intervals. Excluding methane feedstock costs, the estimated abatement cost of ATR-CCS reduces to 172–271 €/tCO_2, making it the most viable short-term option, provided that CO_2 transport and storage facilities are available. This indicates a potential deployment sequence of ATR-CCS, followed by AC or SOEC, depending on site and price conditions.

Four key factors must be considered when determining the timing for deployment. First, recent forecasts suggest an EU ETS allowance price range of 77–160 €/tCO_2 by 2030, followed by a gradual increase until 2039, with a more pronounced rise expected in the early-to-mid 2040s (Homaio, 2024; Mantulet et al., 2023), indicating that economic feasibility for the considered technologies may not be achieved until the late 2030s. However, ambitious emissions reduction targets could prevail over EU ETS allowance price signals as the primary driver of deployment timing. Second, considering planning and construction lead times (3–5 years), deployment must begin early enough for each technology to be operational when EU ETS allowance price levels match their estimated abatement costs. Third, while this timeline allows for the scaling up of critical infrastructure, only three scheduled turnarounds at the cracker plant are expected before 2045, limiting integration opportunities without risking operational disruptions. Finally, policy uncertainties must also be considered, especially concerning the future classification of exported methane as recycled carbon fuel.

Figure 6 illustrates the proposed stepwise deployment strategy aligned with EU ETS allowance price projections and emissions reduction trajectories, as well as turnaround years. Using the steam cracker plant as a representative case, the timeline is divided into short-term, medium-term, and long-term horizons: 2028–2034, 2034–2040, and beyond 2040, respectively. Each phase begins with a scheduled turnaround year to facilitate tie-ins and minimize operational disruptions.

In the short-term, initial efforts should focus on enabling full internal hydrogen recovery from fuel gas[9], allowing export of excess methane to the natural gas grid. This can be achieved through targeted retrofits in the product recovery section or by installing pressure swing adsorbers (PSAs). Importantly, these measures should account for future changes in fuel gas availability due to feedstock switching, cracker decommissioning, electrification, and the integration of emerging recycling technologies (Mandviwala et al., 2024).

In the short-to-medium term (2028–2034), ATR-CCS systems could be deployed due to their relatively low abatement costs (172–271 €/tCO_2), which would be comparable to the projected EU ETS allowance price ranges during this period. However, their deployment will ultimately depend on the availability of CO_2 transport and storage infrastructure. The installed capacities would have to be determined based on the trade-off between cost savings from methane exports and emissions reductions achieved through ATR-CCS during this period. In general, larger ATR-CCS systems are more favorable, as they benefit from economies of scale and can better accommodate the anticipated increase in on-site fuel gas availability with cracker electrification or plastic waste recycling.

The medium term (2034–2040) is a critical period, as it includes the final scheduled turnaround in 2040 before the projected increases in EU ETS allowance prices. Key technologies must, therefore, be deployed and fully operational by this point. Their selection will largely depend on the prevailing market conditions, the availability of renewable electricity, and sufficient grid transmission capacity. In principle, if cheap renewable electricity with sufficient transmission capacity is available, direct electrification of cracker furnaces should be prioritized. However, if direct electrification of existing crackers proves technically unfeasible, or if electricity and adequate grid capacity remain limited, ammonia as a cracker fuel or as feedstock for AC may become indispensable during this period. Moreover, the possibility of using floating facilities for ammonia cracking avoids the use of valuable on-site space and mitigates technology lock-in.

Beyond 2040, the focus shifts toward the installation of electrolyzers as part of an integrated system. In this regard, early installation of electrolyzers would allow for longer operation alongside ATR-CCS and AC, enabling the synergies demonstrated in this work. Given the limitations related to the modular scaling up of electrolyzer systems, investment in overcapacity may be required. Nevertheless, integrated systems with overcapacity tend to offer greater flexibility in responding to price fluctuations while maintaining high capacity utilization. Alternatively, this deployment could be phased, initially targeting the remaining hydrogen demand at the cracker plant and subsequently expanding to meet the green hydrogen demand across the cluster. This would ensure the future hydrogen supply within the cluster, facilitating CO_2 utilization and recycling pathways toward the circular use of materials (Thunman et al., 2019).

In this context, large-scale deployment of low-emissions hydrogen should be pursued as part of a coordinated, cluster-wide effort, rather than as an isolated, plant-level effort. The cluster-wide effort should focus on decarbonizing the steam cracker plant in the short-to-mid-term, as reducing direct on-site emissions (Scope 1) lowers the emissions intensity of intermediate products (olefins) and, in turn, the final products of downstream industries. Thereafter, low-emissions hydrogen could be redirected to preferred end-uses, such as the production of chemicals, as substitute technologies for the steam cracking process become available over time. Moreover, a coordinated approach would help to minimize the risk of stranded assets and enable investment risks to be shared across the cluster, and even outside the cluster via new or repurposed pipelines.

7 Conclusion

This work introduces an integrated hydrogen production system combining autothermal reforming with carbon capture and storage (ATR-CCS), solid oxide electrolysis cell (SOEC), and ammonia cracking (AC) for low-emissions hydrogen supply in petrochemical clusters, and evaluates how technology diversification helps overcome deployment constraints and enhance system flexibility. A methodological framework combining process modeling, integration, and a multi-period mixed-integer linear programming (MILP) model was developed to identify cost-optimal system configurations and dispatch strategies under site-specific constraints and energy market uncertainties. The framework was applied to an existing steam cracker plant, demonstrating that such facilities are a practical starting point for scaling low-emissions hydrogen production in petrochemical clusters.

In this context, several advantages are identified: i) using low-emissions hydrogen as a cracker fuel enables cost reductions through the export of recovered methane from displaced fuel gas; ii) this approach helps to circumvent the absence of both hydrogen and CO_2 transport and storage infrastructure in the short-to-medium term; iii) it avoids the need for costly hydrogen compression and storage systems by using hydrogen directly as it is produced; and iv) it allows hydrogen to be redirected to end-uses with higher CO_2 abatement potential as substitute technologies, such as electrified crackers, become available.

Integrating multiple hydrogen production technologies confers significant advantages over standalone systems in terms of economic performance, operational flexibility, and system redundancy. The results also highlight the importance of securing price certainty through low-cost, long-term contracts for system configurations dominated by a single technology. In the absence of price certainty, cost escalation risks can be mitigated by installing comparable or excess capacities across technologies, enabling cost reductions through flexible operation and hydrogen exports to the cluster.

More specifically, the case study shows that: i) exporting methane recovered from displaced fuel gas reduces on-site

9 A common assumption made throughout the scenario analysis.

hydrogen supply cost (OHSC) by 1–3 €/kgH_2. With these offsets, the OHSC of standalone systems ranges from 3.1 to 3.9 €/kgH_2 (SOEC) and 2.9–5.4 €/kgH_2 (AC), while ATR-CCS achieves lower costs of 1.4–2.3 €/kgH_2 by repurposing methane previously used as cracker fuel into feedstock for hydrogen production; ii) optimal system configurations are highly sensitive to site and price conditions, with levelized costs ranging from 3.0 to 3.7 €/kgH_2 under low-price conditions to 5.6–7.9 €/kgH_2 under high-price conditions. Including methane export offsets, the OHSC ranges between 2.0 and 3.1 €/kgH_2 and 1.2–5.7 €/kgH_2 under low- and high price conditions, respectively; iii) some of the lowest OHSC occur when adequate grid transmission capacity enables SOEC-dominated configurations that achieve significant cost offsets through methane export. In contrast, the highest OHSC occur for AC-dominated configurations under constrained grid and fuel gas availability combined with high ammonia prices; iv) favorable conditions for deployment of each technology emerge at different EU ETS allowance price levels, with ATR-CCS (172–271 €/tCO_2) requiring significantly lower prices than SOEC (373–462 €/tCO_2) and AC (348–647 €/tCO_2), which, even with methane export offsets, require prohibitively high EU ETS prices as standalone systems.

Based on these results, a stepwise deployment strategy is proposed for achieving cost-effective and timely deployment of individual technologies, accounting for key constraints, such as scheduled turnarounds, availability of critical infrastructure, and projected EU ETS allowance prices. Future work could adapt the generalized optimization framework to other clusters, accounting for their site-specific constraints, to support critical infrastructure planning, including CO_2 and hydrogen transport networks and electricity grids. The framework could also be extended beyond hydrogen firing in existing cracker furnaces to include e-crackers and thermochemical recycling technologies toward increased process electrification and circularity. Furthermore, the proposed stepwise deployment strategy could be further assessed under uncertainty using a multistage stochastic programming approach, incorporating technology readiness, infrastructure availability, and future cost developments. This would support investment planning in petrochemical clusters, guiding the transition toward carbon neutrality and circularity enabled by shared hydrogen infrastructure.

Data availability statement

The original contributions presented in the study are included in the article/Supplementary Material; further inquiries can be directed to the corresponding author.

Author contributions

TR: Conceptualization, Data curation, Methodology, Validation, Visualization, Writing – original draft, Software. JB: Methodology, Writing – review and editing. VM: Writing – review and editing. LP: Writing – review and editing. SH: Funding acquisition, Project administration, Supervision, Writing – review and editing. HT: Conceptualization, Funding acquisition, Project administration, Supervision, Writing – review and editing.

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Conflict of interest

The following authors declare financial interests and/or personal relationships that may be considered potential competing interests. Author VM reports employment with Borealis Polyolefine GmbH. Author LP reports previous employment with Borealis AB.

The perspectives, interpretations, and conclusions expressed by authors VM and LP in this scientific article are their personal views and do not necessarily represent the opinions or positions of their affiliated institutions, Borealis Polyolefine GmbH and Borealis AB.

The remaining author(s) declare that this work was conducted in the absence of any commercial or financial relationships that could be construed as a potential conflict of interest.

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Supplementary material

The Supplementary Material for this article can be found online at: https://www.frontiersin.org/articles/10.3389/fceng.2026.1837323/full#supplementary-material

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Nomenclature

AC	ammonia cracker
ATR	autothermal reforming
CAC	CO_2 abatement cost
CAPEX	capital expenditure
CCS	carbon capture and storage
GHG	greenhouse gas
LCOH	levelized cost of hydrogen
LNG	liquefied natural gas
MILP	mixed integer linear programming
MP	medium pressure
OHSC	on-site hydrogen supply cost
PPA	power purchase agreement
SMR	steam methane reformer
SOEC	solid oxide electrolysis cell