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Techno-Economic Analysis of Small-Scale Electro-Ammonia Production in a Port Platform for Maritime Transport

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Abstract

Maritime transport is energy-efficient but remains heavily dependent on fossil fuels. Renewable electricity–based ammonia (e-NH_3) has emerged as a promising alternative, particularly through small-scale, modular production. Assessing its economic viability is essential for future adoption, and techno-economic analysis offers a structured way to evaluate its feasibility. This study investigates the cost performance of a small-scale offshore e-NH_3 plant of 2.4 tpd at the Port of Santander, Spain, based on nitrogen obtained via membrane separation and hydrogen from electrolysis of pretreated seawater. The results include process simulation outcomes obtained with ASPEN v14, a detailed cost breakdown based on modular costing methodologies applied to preliminary process designs, and sensitivity analyses of the levelized cost of e-NH_3 (LCOA) with respect to the variables that have the strongest influence on overall costs. A comparative review of LCOA values reported in the literature for offshore and onshore e-NH_3 plants is provided. An estimated CAPEX of 5.99 M€ (equivalent to 0.53 M€/y), OPEX of 1.58 M€/y, and a LCOA of 2408 €/tNH_3 are obtained, with equipment investment and operating costs identified as the most influential parameters. The results highlight the need for supraregional techno-economic studies, considering optimal offshore wind availability within a collaborative interregional framework.

Keywords: ammonia; e-NH_3; small scale; green hydrogen; renewable energy; techno-economic assessment; offshore; LCOA

1. Introduction

Maritime transport plays a fundamental role in the global economy and remains one of the most energy-efficient modes of transportation. However, the sector is still almost entirely dependent on fossil fuels. In the International Energy Agency's (IEA) Net Zero Emissions by 2050 Scenario, the use of renewable hydrogen and electrofuels (e-fuels) in the energy sector, primarily in transport, shows significant medium- to long-term growth potential, with projected expansion from near-zero levels today to 1.5 exajoules by 2030 [1].

The 2023 strategy of the International Maritime Organization (IMO) outlines several "levels of ambition," including: (i) reducing CO_2 emissions per transport work by at least 40% by 2030 compared to 2008 levels, and (ii) achieving net-zero greenhouse gas (GHG) emissions from international shipping by 2050, among other goals. Technological innovation and global adoption of alternative fuels and/or energy sources for international maritime transport will be essential to meeting these ambitions [2]. In parallel, the European Union's FuelEU Maritime regulation aims to increase the share of renewable and low-carbon fuels in the maritime energy mix. This proposal seeks to reduce GHG emissions on board vessels up to 75% by 2050, thereby promoting the use of more sustainable fuels within the shipping sector [3].

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In this context, hydrogen has been proposed as an energy carrier, that can be produced in regions with abundant renewable resources and transported to areas where needed [4–6]. Nevertheless, the use of hydrogen presents significant challenges, including low volumetric energy density and complex conditions required for storage and transport. In order to overcome these limitations, other hydrogen-derived compounds have been proposed within the framework of Power-to-X (PtX) processes. These involve the conversion of renewable electricity into chemical products or e-fuels, known as Power-to-Fuel (PtF). Among these products, ammonia stands out due to its potential to serve as: (i) an efficient energy carrier, (ii) a medium for electricity storage, (iii) a fuel for various applications such as maritime transport, and (iv) a feedstock for the chemical industry.

Ammonia represents a viable carbon-free, non-fossil fuel alternative for maritime transport over the medium to long term. Solid Oxide Fuel Cells (SOFCs), as well as hybrid systems combining SOFCs with reciprocating engine generators, have shown considerable potential for near-term deployment in marine applications [7,8]. A major advantage of ammonia lies in its well-established infrastructure for production, storage, and distribution, owing to its long-standing use in the fertilizer industry. Nonetheless, the transition to green ammonia presents several challenges, including compliance with safety regulations, engine adaptation (e.g., dual-fuel configurations), scale-up/down capacity, and continued innovation in renewable energy technologies. Although such challenges are comparable to those faced by other emerging green fuels, proactive measures are essential to address these barriers and enable its broader adoption [9].

Several authors have conducted environmental impact studies and life cycle analyses on the use of e-NH_3 as a marine fuel [10–13]. Although green ammonia may present a lower environmental impact compared to other fuels, these studies may show discrepancies depending on the methodology, case study, scope, impact categories, analysis boundaries, data, and assumptions employed. From a regulatory perspective, a significant barrier to the use of ammonia as a fuel may arise from future legal regulations in the maritime sector. In this regard, the IMO [14] is expected to approve interim guidelines for ammonia fuel use, providing the shipping industry with its first international standard for safe ammonia fuel operations.

One of the key aspects to consider for the future adoption of e-NH_3 in maritime transport is its economic viability, given the significant economic risks associated with this fuel, as highlighted by Fullonton et al. [15]. Additionally, several factors have been identified as major challenges for the integration of green ammonia into the maritime supply chain, including safety, ammonia propulsion technology, fuel availability, uncertain regulatory framework, and intermittency and variability of renewable electricity generation from sources like solar and wind [16,17]. Despite the remaining challenges, the prospects for the production and use of e-ammonia as a maritime fuel may contribute to the achievement of the Sustainable Development Goals (SDGs), particularly SDG 7 (Affordable and Clean Energy) and SDG 13 (Climate Action), as it relies on renewable energy sources, atmospheric nitrogen, and water. Moreover, projections indicate that green ammonia could power up to 99% of the marine sector by 2050, underscoring its relevance to SDG9 (Industry, Innovation, and Infrastructure) [18–20].

Currently, there is no fully mature application of liquid ammonia carriers in ship transportation [21]. However, several authors have explored the use of ammonia as a marine fuel. Inal et al. [22] provided evidence that ammonia is the most suitable zero-carbon fuel for the application of the marine fuel cell. The review by Machaj et al. [23] discusses ammonia from technological, scientific, economic, and regulatory perspectives, presenting a technical justification for using solid oxide fuel cells to enhance energy conversion efficiency. In Braun et al. [24], the feasibility of using ammonia as a green fuel for inland waterway vessels is assessed, highlighting onboard hydrogen release, ammonia-hydrogen combustion in Internal Combustion Engines (ICEs), and thermal integration. The study concludes that small- and medium-sized ships can operate effectively with proper system integration. Wang Z et al. [25] propose a green ship power system based on H_2 and NH_3 using a "demand-configuration-integration-evaluation" framework, identifying six key requirements for zero-carbon fuel ships: emissions, power, cost, safety, volume, and noise. Caprace et al.[26] reveals that shipowners prioritize biofuels while higher-cost alternatives like ammonia remain economically unviable without additional policy support. Wang and Li [27] shows that ammonia-based fuels serve as transitional solutions to adopt sustainable transportation solutions showcasing their long-term potential for zero-emission shipping. Currently, there are few examples of commercial applications for ammonia on ships, such as the Norwegian Viking Energy and the Scandinavian MS Green Ammonia; however, several ammonia-powered and dual fuel vessels are expected to enter operation in 2026, following advances made across the ammonia fuel value chain [17,28].

Supportive policies for demonstration projects, fundamental R&D, and case study analyses will help reduce cost-related barriers and facilitate the implementation of regulations that ensure a level playing field for e-fuels in comparison with conventional alternatives. Techno-economic analysis applied to e-ammonia production methods enables the assessment of their technical readiness and economic viability, and can identify key technical and financial barriers to its implementation. In particular, small-scale production, along with modularity and mild operating conditions, are of special interest for the sustainable production of e-ammonia. This modular approach allows for progressive, flexible, and decentralized development of production units, reducing technological and economic risks while facilitating integration with renewable energy sources. Furthermore, small-scale analysis enables experimental validation and controlled scaling, which are key factors for accelerating the industrial adoption of sustainable solutions within the framework of global decarbonization.

The wide range of potential applications for e-NH_3 is currently driving the planning of numerous production projects linked to the generation of e-H_2 from renewable energy sources. Of particular interest is the use of e-NH_3 for heavy-duty, long-haul segments in the maritime transport sector [29,30]. In this context, both at the Spanish national level and within the Cantabria region (Northern Spain on the coast of the Cantabrian Sea), various activities related to the marine energy value chain are being developed [31,32]. Nevertheless, techno-economic studies on small-scale offshore e-NH_3 production remain relatively scarce.

The main purpose of this study is to assess the costs associated with the small-scale offshore production of e-NH_3 at a rate of 2.4 tpd (100 kg/h), based on nitrogen obtained via membrane separation and hydrogen produced through electrolysis. All the processing units are located on an offshore platform in the Port of Santander (Cantabria, Spain). This paper is structured as follows. Section 2 presents a literature review of e-NH_3 production costs, with a particular focus on offshore systems, in order to compare the plant under study with other production scales and locations. Section 3 provides a process description within the defined system boundaries, focusing on the primary process. The methodology in section 4 outlines the technical analysis, equipment sizing, and cost estimation approach; it also presents the cost allocation criteria and the main technical and economic assumptions. The results and discussion in section 5, presents the outcomes of the analysis, including process simulation results, a detailed cost breakdown, and sensitivity analyses of the levelized cost of ammonia (LCOA) with respect to the variables exerting the greatest influence on overall costs. Finally, the results are summarized and conclusions are derived in Section 6. The findings aim to support policymakers in promoting e-ammonia as a sustainable energy solution for maritime transport.

2. Literature review Of e-NH_3 Production Cost Assessment

A comparative overview of research works related to green ammonia production plants has been conducted for both offshore (Table 1a) and onshore (Table 1b), organized by increasing production capacity. This literature analysis includes production capacity (in tons per day, *tpd*), Capital Expenditure (CAPEX), electricity cost, Operational Expenditure (OPEX), and the Levelized Cost of Ammonia (LCOA), as well as contextual details and references. To ensure comparability, the economic data extracted from the different studies were first updated to constant 2024 prices to account for inflation, and subsequently converted into euros.

Firstly, offshore plants typically show lower production capacities, with many plants producing below 300 tpd, though one large-scale scenario reach 2,740 tpd [33]. On the contrary, onshore plants include a wider range of scales, from small pilot plants (<10 tpd) to industrial-scale facilities (>3,000 tpd). This suggests that onshore plants are currently more scalable.

Regarding the costs, it can be observed that, in both offshore and onshore cases, CAPEX per tpd tends to decrease with increasing plant scale, although there is a wide variability depending on the technology and the location. At very large scales (approaching 3,000 tpd), CAPEX decreases more, suggesting that huge scale production is the most cost-effective configuration. Offshore plants CAPEX is, in general, higher per unit of production than onshore plants CAPEX. The need for specialized infrastructure, such as wind turbines or offshore platforms, significantly increases capital costs.

Small-scale demonstrators or pilot plants exhibit high capital costs per ton of daily ammonia. At very low capacities (<20 tpd), CAPEX can be as high as that of large-capacity plants. This is due to the fact that infrastructure, equipment, and other systems are often similar in complexity and cost to larger installations, but are amortized over a much smaller volume of production. The small-scale plants are typically early-stage or experimental projects and not optimized for economic performance.

In relation to operational expenditure (OPEX), smaller installations, particularly those offshore or in remote locations, tend to have a higher relative OPEX, due to operational complexity, maintenance, and logistical costs. In contrast, larger plants, especially those with stable energy supply and infrastructure (typically onshore), have lower OPEX percentages. This is consistent with the fact that larger systems spread fixed operational costs over a higher production volume. For instance, some large-scale systems explicitly report OPEX as low as 2% of CAPEX per year [34].

These relationships between plant scale, CAPEX and OPEX are reflected in the Levelized Cost of Ammonia (LCOA). Plants with low production capacity and high unit CAPEX/OPEX report high values of LCOAs, around 1,000 – 4,000 €/tNH_3, which are economically uncompetitive when compared with typical of conventional large-scale (260–500 €/tNH_3) or low scale (≃520 - 1,000 €/tNH_3 applying six tenths rule to scale down) grey ammonia production based on natural gas reforming [34–36].

It's also worth noting that electricity cost is included in OPEX calculation and it is a major cost driver in LCOA. Offshore production heavily depends on wind or hybrid renewable sources; therefore, site selection with favourable wind profiles and shorter distances to shore can reduce the values of LCOA. In optimal conditions (e.g., strong wind or proximity to grid), LCOA could potentially reach levels below 800 €/tNH_3, approaching the cost of grey ammonia in some markets [37].

Although the comparability of the data presented in Tables 1a and 1b may be affected by differences in the assumptions adopted across the studies—such as system boundaries, energy sources, process units, and technology types—the data reflects that green ammonia production is becoming more developed, especially for onshore plants, which are currently more cost-effective and easier to scale up. Offshore plants have potential, especially for energy integration and decentralized production, but they still face high costs and technical challenges that must be addressed for wider adoption.

A growing body of additional techno-economic studies, not included in Tables 1a and 1b, conducts comparative analyses of the global potential for green ammonia production. These studies examine a large number of locations and countries in diverse geographic regions worldwide, addressing a large number of scenarios that provide a huge amount of data beyond of the scope of Tables 1a and 1b [33,38–42]. They are framed within a global context characterized by substantial renewable energy potential, an urgent need to mitigate CO_2 emissions, and a growing worldwide demand for both energy and fertilizers. In this context, e-NH_3 emerges as a promising energy carrier and a key chemical feedstock for sustainable fertilizer production. These analyses explore various renewable energy sources and system configurations encompassing production, transportation, and distribution, adopting a full supply chain perspective to identify geographically favourable locations and to project production costs up to 2050. Although projected LCOA vary widely, within the range of approximately 2,000–260 €/tNH_3, several future scenarios indicate that e-NH_3 could achieve cost competitiveness with conventional fossil-based ammonia.

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Table 1. a. Comparative overview of e-NH_3 production offshore plants cost assessment. (Monetary units expressed in updated 2024 euros).

NH_3 Production (tpd)	CAPEX (M€) *(Full plant)*	Electricity cost (€/kWh)	OPEX (M€/y)	LCOA (€/t_{NH3})	Comments (Cost basis year. Location)	Ref.
2.4	5.99	0.0199 *(Wind)*	1.58	2,408 (2024)	2024. Spain. Port of Santander. *(100kg/h)*	Present work
15	20.1	0.122	4.7	1,821	2021. Nova Scotia, Canada.	[43]
50 – 1,500	33.6–44.8 equipment 11.2–30.3 platform 279–314 turbines 112 offshore platform	0.022	22.4 equipment 52.7 turbines	784-2,128	2020. Various scenarios considering wind profiles, ammonia demands, distances to shore, and water depths.	[37]
54.8	36-38	0.011-0.022 *(Wind)* *0.046* *(Mixed)*	17.5-19.4	1,045-1,154	2020.HB synthesis with two alternatives. *(20.000 t/y)*	[44]
274	784 (WT) 896 (W)	----	2%CAPEX	903 (WT), 1,028 (W)	2020. Pentland Firth, Scotland Wind(W) + Tidal (WT)	
300	1848	0.115	---	1,628	2010. USA. Gulf of Maine. Costs includes NH_3 facility + wind farm	[45]
300 max.	462-707	---	---	1,813	2018.USA. 65-100 % loads.	[46]
2,740	242 for offshore platform	---	---	1,344	2020.Global offshore production locations. *(1 MMtpa)*	[47]

Table 1. b. Comparative overview of e-NH_3 production onshore plants cost assessment. (Monetary units expressed in updated 2024 euros).

NH_3 Production (tpd)	CAPEX (M€) *(Full plant)*	Electricity cost (€/kWh)	OPEX (M€/y)	LCOA (€/t_{NH3})	Comments (Cost basis year. Location)	Ref.
2.74	9.87	0.0517	0.25	1,038	2025. South of Sardinia, Italy. Scenario without energy revenue	[48]
4	80	0.091	0.94	4,317	2025. Morocco.	[49]
8.7	----	0.028-0.072	----	1,009-1,801	2020. UK. Scenarios with battery storage, H_2, or flexible HB.	[50]
27.1	70.2	----	17.66	1,030	2020. Values with maximum production.	[51]
68.5	66.8	0.045	30.65	1,398	2020. Norway. Hydroelectric energy.	[52]
95.9	165.8	0.038 – 0.051	3% CAPEX	518 – 795	2017&2020. Chile and Argentina.	[53]
100 - 61	Several references by unit	0.135	Several references by unit	850 (2030) 540 (2050)	2030 and 2050 scenario. Italy. Mix of wind and	[54]

					solar power. 20-100% flexibility.	
137 – 2,740	49-681	----	2-5% CAPEX	560-1,288	2019&2020. Onsite and coastal scenarios.	[40]
137.04	215.5	0.045	37.4	1,134	2019. France	[55]
240	373	0.022	7,45	675	2025 forecast. Islanded production	[56]
273.97	530.43-1,057.15	0.0599 *(Grid)*	59.67 – 21.97	925 – 1,163	2024. South Africa. Three electrical scenarios.	[57]
274	Electrolyser: 39.6 H_2 storage: 8,.9 HB: 55.6	0.056	1-2% CAPEX	543.9	2022. China.	[58]
300	1,085-1,414	0.24	67-108	1,040-1,500	2019. Germany. Alkaline and PEM Electrolysers	[59]
680	168	----	23.6	----	2023. USA.	[60]
1,178	Cost of plant components depending on production	0.03	176	710 (2020) 660 (2030) 550 (2050)	Applicable to an economic analysis for a specific site	[61]
1,840	---	0.028	2%CAPEX	805	(2020). Regions with insolation. Aspen Economic Analyzer.	[34]
2,378	1,192	0.040	672	1,081	2022. Australia	[62]
2,400	2,313	0.011-0.045	72.8 (Fixed)	588-1,311	2020.	[1]
3,000	---	0.06	---	676.9	2020. Spain	[35]
3,000	----	0.043 (Solar&wind) 0.047 *(Grid)*	158	338.5	2019. Australia	[63]
3,300	1,366	0 – 0.098	---	0 – 980	2018. Texas, USA	[64]

3. Process Description

The e-ammonia production plant under study comprises four sequential units: seawater purification, H_2 generation, air separation, and NH_3 synthesis (Figure 1). The power supply required for the plant will be provided by offshore wind energy. Due to the intermittency of renewable energy supply, storage systems for H_2 and electricity are required. The plant is designed for a production capacity of 100 kg/h (2.4 tpd) of ammonia with a minimum purity of 99.6%, using a 1.5 MW electrolyser. The diverse techno-economic studies on e-ammonia production presented in Tables 1a and 1b highlight that the economic assessment of this process is site-specific. Therefore, the Port of Santander in Cantabria Region, Spain has been selected as a case study for the development of the techno-economic analysis.

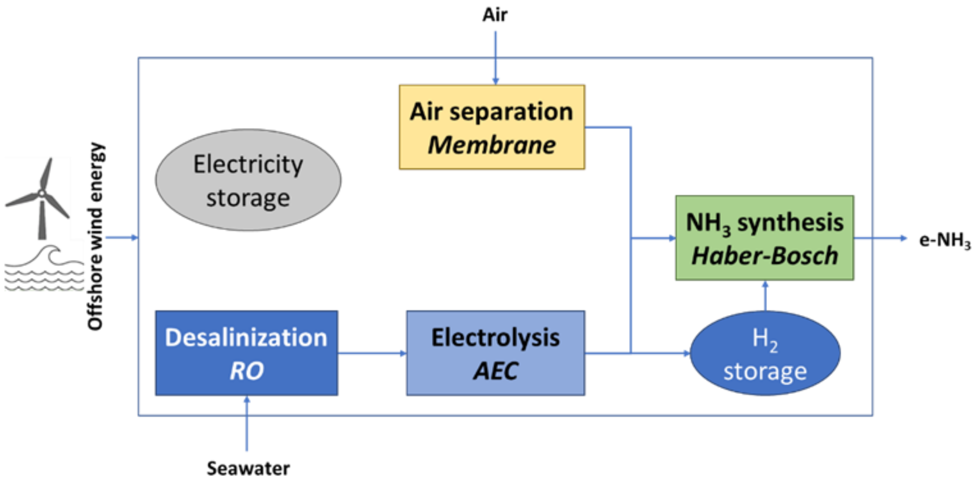

Figure 1. Boundary of the studied e-ammonia plant.

The supply of purified water is ensured through a reverse osmosis (RO) unit, a technology characterized by relatively low energy consumption and cost, capable of providing water with a total dissolved solids (TDS) content of less than 3 ppm and conductivity below 5 µS/cm, as required for alkaline electrolysis [65,66]. The unit consists of pretreatment via Dissolved Air Flotation (DAF) coupled with a stage of dual media filtration, high-pressure pumps, and a membrane array housed in pressure vessels. DAF equipment is an effective pretreatment system for seawater desalination, commonly employed as water pretreatment for reverse osmosis (RO) desalination systems [67–69]. The DAF system, when combined with filtration, is particularly efficient in removing algal cells, oil and grease, and suspended solids present in seawater. Reported removal efficiencies range from 90–98% for suspended solids and exceed 99% for algal removal [70–72].

The air separation unit (ASU), which provides the N_2 required for synthesis, is based on membrane separation technology, the preferred alternative for small-scale systems [73,74]. An optimal configuration is proposed using three membrane modules to achieve a nitrogen purity of 99.9% [75,76]. In addition to the membrane modules, the main components of the ASU include compressors, vacuum pumps, and heat exchangers.

Alkaline electrolysis (AEL) is the selected technology in this study for hydrogen production [5,55], yielding hydrogen at 99.5% purity and a pressure of 30 bar. Alkaline electrolyser stacks enable flexible operation, with the ability to rapidly ramp up or down and operate across a wide range of conditions. These characteristics are essential for effectively managing the variability and intermittency associated with renewable energy sources. In addition to the alkaline electrolyser cell, the unit includes pumps, heat exchangers, and flash equipment.

The synthesis unit is based on the Haber-Bosch process and includes compressors, reactors, heat exchangers, a synthesis loop with recycle and purge streams, and flash equipment. The unit is designed with a configuration of two reactors in series operating at pressures between 205 and 250 bar [77]. The produced ammonia would be stored at the Port of Santander facilities, in parallel with the existing liquefied natural gas storage infrastructure used for current maritime transport.

Gaseous hydrogen storage is presently the most widely employed technology for storing hydrogen. Medium-to-low-pressure storage aligns well with the pressure of hydrogen gas generated by the electrolyser [58]. For this study, a 3 MPa low-pressure hydrogen storage spherical tank is chosen as the hydrogen storage equipment. For energy storage, a fully integrated modular system consisting of rechargeable lithium-ion batteries with high energy density, long service life, and high efficiency is selected, intended for wind energy integration. The system can operate as a microgrid to support backup and islanded operations. All described units will be located on an offshore platform situated in the Port of Santander, Spain.

In Figure 2, the process diagram is shown, with the four process units, distinguished by different colors, with all the equipment included.

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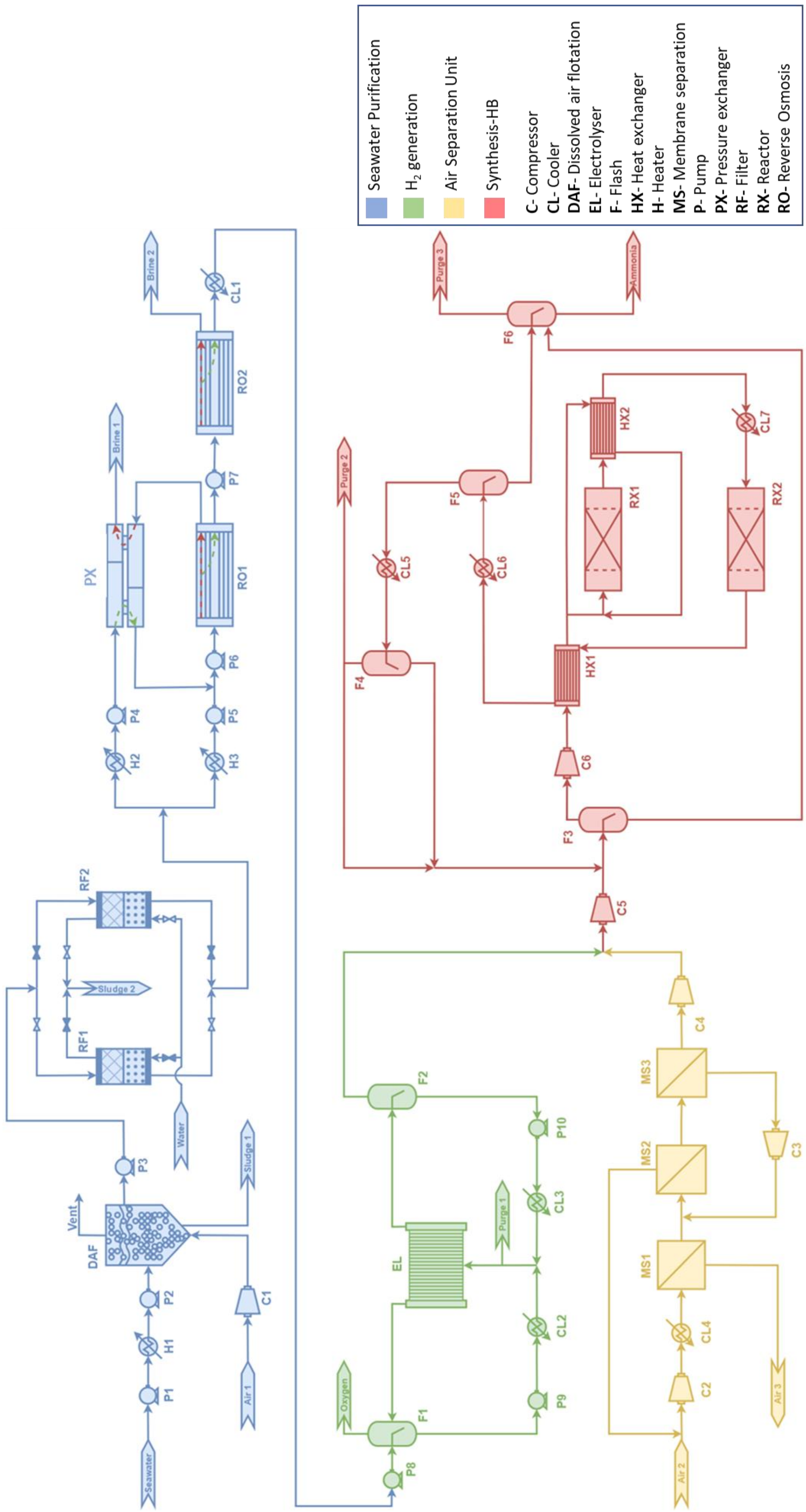

Figure 2. Process Flow Diagram of the studied e-NH3 plant.

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4. Methodology

The applied methodology follows a structured approach for estimating the costs of the ammonia production plant, integrating process simulation, equipment sizing & costing, and Capital & Operating Expenditures estimation. The process involves different key steps, from mass and energy balances analysis to the final estimation of the Levelized Cost of Ammonia (LCOA).

The four units that comprise the e-ammonia production plant are steady-state simulated using Aspen Plus® Version 14 software. The process simulation is based on a previous work of the research team [77], incorporating a seawater desalination unit and implementing heat integration within the Haber–Bosch synthesis unit. The process analysis enables the determination of the mass and energy balances, as well as the operating conditions, which is the basis for equipment sizing and costs estimating.

The design of the various units across the process flowsheet—such as storage tanks, flash separators, and heat exchangers—is carried out based on standard design methodologies described by Sinnott and Towler [78] and Biegler et al. [79]. Pumps and compressors are sized according to their power requirements, while the Haber–Bosch (HB) reactors are designed based on kinetic models available in the literature [34,80]. The sizing of the desalination unit (including pretreatment and reverse osmosis) and the air separation unit (ASU, using membrane modules) is performed by specific design correlations [77,81].

For preliminary equipment cost estimation, the Guthrie modular method is employed to account for both direct and indirect costs. This module costing technique is particularly suitable for early-stage feasibility studies and is widely adopted in chemical process design due to its flexibility and ease of use. This method yields the updated Bare Module Cost (BMC), which incorporates correction factors for material type, operating pressure, equipment complexity, installation costs, and monetary update to 2024 values [79]. In addition, equipment cost correlations provided by Seider et al. [82] are applied. A detailed breakdown of the formula used to calculate the BMC for each equipment type is provided in the Table S1 of the Supporting Information.

The cost estimation of the electrolyser, the storage systems and the platform were carried out through a dedicated literature review described in Section 5 of the present work.

4.1. Capital expenditures (CAPEX) and operating expenditures (OPEX)

The capital expenditures (CAPEX) of the designed plant include the Total Permanent Investment (TPI) and the Working Capital (WC). The TPI is calculated as a function of the BMC and the platform acquisition cost and encompasses all investments required for plant construction and commissioning, including equipment procurement, installation, and project development. The Working Capital has been estimated at 5% of the TPI for a simple, single-product process without product storage, following the guidelines of Sinnott and Towler [78]. The commonly used Capital Recovery Factor (Eq. 2) allows the annualization of CAPEX as a function of the annual interest rate (i) and the plant lifetime (PL) [83].

$$\mathbf{CAPEX} = C_{TPI} + C_{WC} = (1.18 \cdot \sum BMC + C_{platform}) + C_{WC} \quad (1)$$

$$\mathbf{ANNUALIZED\ CAPEX} = \frac{i\,(1+i)^{PL}}{(1+i)^{PL}-1} \times CAPEX \quad (2)$$

The estimation of operating expenditures (OPEX) accounts for yearly operating costs and comprises the sum of direct manufacturing costs (DMC), fixed manufacturing costs (FMC), and general expenses (GE). Each of these components can be estimated using the updated Bare Module Cost (BMC), cost of operating labor (C_{OL}), cost of utilities (C_{UT}), cost of waste treatment (C_{WT}), and cost of raw materials (C_{RM}), by applying appropriate multiplication factors as described in Equation 3 [83].

$$\mathbf{OPEX} = DMC + FMC + GE = 0.26 \cdot \sum BMC + 2.18 \cdot C_{OL} + 1.075 \cdot (C_{UT} + \cancelto{0}{C_{WT}} + \cancelto{0}{C_{RM}}) \quad (3)$$

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For the estimation of operating labor, the rule-of-thumb presented in Equation 4 has been applied, where P represents the number of processing steps involving solids handling—zero in this case—and Nnp is the number of non-particulate processing steps. The operator salary has been sourced from the 21st General Collective Agreement for the Chemical Industry in Spain for the year 2024 [84]. The utility cost in this study includes the electricity consumption of all electrically driven equipment. An average electricity price of 0.0199 €/kWh has been assumed, corresponding to the mean wind energy price in Spain in 2024, compared to the average wholesale electricity market price of 0.06303 €/kWh [85]. The waste streams in the electric ammonia plant consist primarily of brine from the reverse osmosis (RO) unit and oxygen and argon streams from air separation and electrolysis units. All waste streams are assumed to be environmentally friendly, and hence, waste treatment costs are considered negligible. The cost of raw materials is not included, as seawater and atmospheric air are used for ammonia production.

$$\text{Number of operators} = (6.29 + 31.7 \cdot P^{2} + 0.23 \cdot N_{np})^{0,5} \quad (4$$

4.2. Levelized Cost of Ammonia (LCOA)

Levelized Cost of Ammonia (LCOA) in €/tNH$_3$ is the sum of the present value of the capital expenditures and the operating expenditures over the lifetime of the system, divided by the total ammonia production (Eq.5). This cost metric enables direct comparison with values reported in scientific literature, providing a benchmark for economic feasibility.

$$\mathbf{LCOA} = \frac{\frac{i\,(1+i)^{PL}}{(1+i)^{PL}-1} \times \text{CAPEX} + \text{OPEX}}{F} \quad (5$$

Where *i*, is the discount rate (%), PL the Plant lifetime (years) and F is the yearly ammonia production (t/year)

The main economic assumptions regarding capital and operating cost estimation and LCOA are presented in Table 2.

Table 2. Economic evaluation assumptions.

Target cost basis	Assumption
Location	Spain
Plant capacity	2.4 tpd
Year	2024
Currency	€
CAPEX estimation	
Bare Module Cost (BMC)	(UF) x (BC) x (MF+MPF-1)
Update Factor (UF)	CEPCI 2024/CEPCI Base = 799.1/115
Bare Cost (BC), Module Factor (MF) and Material & Pressure Factors (MPF) (Correction Factors)	Based on Williams rule of thumb (economy of scale), Guthrie modular method and equipment cost correlations provided by Seider et al. [82]
Capital expenditures (CAPEX)	CTPI + Cwc
Total Permanent Investment (CTPI)	1.18 (ΣBMC) + Cplatform
Working capital (CWC)	5% of the CTPI
Electrolyser cost	750 €/kW
Platform cost	2 M€
Lithium-ion battery racks cost	100 €/kWh
Hydrogen storage tank	350 €/kg

OPEX estimation	
Operation expenditures (OPEX)	OPEX = DMC + FMC + GE = 0.26 ΣBMC + 2.18 COL+ 1.075 (CUT+CWT+CRM)
Operating labour unitary cost (COL)	28,143.06 €/y
Wind energy unitary cost	0.0199 €/kWh
LCOA	
Discount rate	8%
Plant lifetime	30 years

5. Results and Discussion

This section presents the simulation, sizing and costing results of the 2.4 tpd of e-ammonia production plant, assuming its location at the Port of Santander, in Cantabria region, Spain. All economic values in this study are expressed in 2024 euros.

5.1. Process Analysis: Mass & Energy Flows and Operating Conditions

The process simulation yields the mass and energy balances, the feedstock and utility requirements, as well as the operating conditions necessary for equipment sizing. The input materials for the production process include 544 kg/h of seawater supplied to the reverse osmosis (RO) unit and 120.5 kg/h of air to the air separation unit (ASU). Additionally, 14.4 kg/h of compressed air is required for the dissolved air flotation (DAF) pretreatment system.

The simulation of the desalination unit consists of a dissolved air flotation (DAF) system, filtration, and reverse osmosis (RO). The membrane module for desalination was simulated in Aspen Custom Modeler (ACM) and subsequently exported to Aspen Plus to construct a flowsheet model of a two-stage seawater desalination system with energy recovery via a pressure exchanger. Seawater is mixed with compressed air in a dissolved air flotation (DAF) unit, where air bubbles injected at 3 bars attach to suspended solids, enabling them to float to the surface. The DAF unit is modelled as a "Flash3" block, which allows representation of a third solid phase. These suspended solids, being a complex mixture of organic and inorganic compounds, were defined as a nonconventional component because their composition is not represented by a simple chemical formula but rather by their elemental analysis and higher heating value (HHV). The user must provide these inputs to Aspen Plus to enable estimation of physical properties such as density and enthalpy. In this work, data corresponding to algae [86] were considered, with a concentration of 20 mg/L. The clarified water is directed to a dual media filtration (DMF) system consisting of anthracite and sand and is subsequently split into two streams. The upper stream is pumped to the pressure exchanger (PX), where it receives energy from the retentate of the first reverse osmosis stage (RO-1); the lower stream is pressurized to match the outlet pressure of the PX, ensuring proper pressure balance in the downstream mixer. The reverse osmosis system, with a total membrane area of 8 m^2, achieves a recovery of 32% in a configuration closely resembling that of commercial reverse osmosis systems [87]. The desalination unit delivers 164 kg/h of water suitable for AEC electrolysis, with a total dissolved solids (TDS) content of 1.5 ppm and an electrical conductivity of 2.8 µS/cm. This is achieved from an input flow of 544 kg/h of seawater with an initial TDS of 35,000 ppm and a conductivity of 71,000 µS/cm [88].

For small-scale systems, membrane permeation is the preferred option for nitrogen production. These systems typically include three membrane modules and compression or vacuum equipment to handle the feed and permeate streams [74–76]. The ASU unit is based on three membrane modules with a total membrane area of 444 m^2. Air is compressed to 2.9 bar and heated before entering the first membrane module (MOD1), producing an oxygen-enriched permeate. The retentate sequentially feeds MOD2 and MOD3, where high-purity nitrogen (99.9% molar) is obtained. The permeates from MOD2 and MOD3 are recompressed and recirculated to the preceding modules, while vacuum pumps maintain the pressure gradient of 0.2 bar required for efficient oxygen permeation. The membrane modules are simulated using the "Sep" block, which enables separation based on the component fluxes. In this work, the fluxes are determined according to the membrane area and the permeability values of commercial membranes [77]. The unit delivers a nitrogen stream of 85 kg/h with a purity of 99.9%.

Considering small-scale systems, alkaline electrolysis was selected in this study as the hydrogen production technology. An aqueous stream containing 35 wt% potassium hydroxide is fed to the alkaline electrolyser, yielding two gaseous output streams: hydrogen at the cathode and oxygen at the anode. The potassium hydroxide solution is continuously recirculated, while the water consumed during electrolysis is replaced. Subsequently, the O_2 and H_2 streams are purified to obtain high-purity gases (>99.5%). The liquid streams from the flash unit are recirculated to the process and a safety purge is implemented to maintain stable feed conditions to the electrolyser. For the "electrolyser" block, a physical model available in Aspen Plus® V14 was employed. The unit was designed based on specifications provided in the technical data sheets of commercial electrolysers. The alkaline electrolysis process produces 18.4 kg/h of hydrogen with a purity of 99.94%, operating at 30 bar and 60 °C.

The Haber–Bosch unit consists of a compressor that pressurizes the synthesis loop, a reactor system for converting H_2 and N_2 gases into NH_3, a heat exchanger that utilizes the exothermic heat of reaction to preheat the reactor feed stream, a condensation-based separation system to extract ammonia from the product stream, and recycle and purge lines to recover unreacted H_2 and N_2 while removing inert gases to prevent accumulation in the loop. In the present work, the reactors have been modelled using an "RPLUG" block in Aspen Plus V14, according to de la Hera et al. [77]. The HB plant design developed in this work includes two heat exchangers between process streams, which allow the reactor inlet streams to reach the optimal reaction temperatures required to maximize conversion while maintaining suitable reaction rates. This heat integration reduces the overall energy demand of the unit, which is mainly concentrated in the coolers located upstream of the ammonia separation systems (Figure 2). The Haber–Bosch unit operates in a dual-reactor configuration [77], producing 100 kg/h of ammonia with a purity of 99.6%. In Table 3, the main material and energy flows obtained are summarized.

Table 3. Main material and energy flows.

Material flows of the process (kg/h)	
Inlet seawater flow	544
Inlet water flow to electrolysis unit	164
Inlet air flow to ASU unit	120.5
Inlet H_2 flow to the synthesis unit	18.4
Inlet N_2 flow to the synthesis unit	85
Outlet ammonia flow	100
Energy requirement of the process (kWh/kg NH_3)	
Full plant	10.55
Desalination unit	0.09
Electrolysis unit	10.0
Air separation unit	0.05
Synthesis-HB unit	0.41

The overall specific energy consumption of the e-ammonia plant is estimated at 10.55 kWh/kg NH_3 of which 94.8% corresponds to hydrogen production, 3.9% to ammonia synthesis, 0.5% to nitrogen production, and 0.9% to seawater desalination, as it can be observed in Figure 3. Under regular operating conditions, the electrolyser alone requires most of the electrical power, which is significantly higher than the consumption of the other process units and clearly dominates the plant's overall energy demand.

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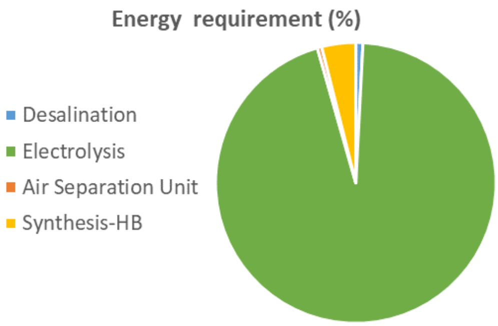

Figure 3. Energy requirements of process units of the studied e-NH_3 plant.

To reduce the overall energy consumption of the plant, two energy integration strategies have been proposed: the use of a pressure exchanger (PX) in the desalination plant and the implementation of heat integration through shell-and-tube heat exchangers (HX1, HX2) in the ammonia synthesis unit. The energy recovery system employed in seawater reverse osmosis is a pressure exchanger (PX), in which hydraulic energy is recovered by transferring pressure from the high-pressure brine reject stream (retentate) to the feed stream, using only one moving part [89]. The equipment cost estimation is based on data reported by Timur et al. [90] and Wang et al. [91], extrapolated to the purified water production levels considered in the present study, and complemented with purchase cost correlations for a liquid expander [82]. The installed equipment cost falls within the range of 3,900–10,000 € depending on the model, efficiency, pressure and local conditions, with a value of 7,000 € adopted in this work for a very small plant without economies of scale. This price corresponds to that of small-sized commercial units available on the market [92].

The heat exchangers in the Haber–Bosch unit enable heat integration between process streams by preheating the inlet streams to the first reaction stage (RX1) using the hot outlet streams from the two reactors; these hot streams are subsequently cooled in HX1 before entering the flash unit (F5) and in HX2 before entering the second reaction stage (RX2).

5.2. Equipment Sizing and Costing

Equipment sizing is the basis step for a reliable economic assessment. The sizing of each unit operation depends on the results of process mass and/or energy balances, depending on the type and function of the equipment. In table S1 of the Supporting Information the equipment sizing parameters are included, organized by each unit of the ammonia production plant. Table S2 of the Supporting information includes the obtained values of the sizing parameters of the different equipment. Based on these parameters, the investment and operating costs of the plant are estimated, as well as the Levelized Cost of Ammonia (LCOA). All prices in this study are expressed in 2024 euros, adjusted for inflation using the average value of 799.1 from the Chemical Engineering Plant Cost Index (CEPCI) of 2024 [93].

The application of the Guthrie modular method for cost estimation, combined with the equipment cost correlations provided by Seider et al. [82], enables the estimation of installed equipment costs (Bare Module Cost, BMC), which are shown in Table S3 of the Supporting Information.

The costs of the electrolyser, the hydrogen and energy storage and the offshore platform are analysed separately and discussed in the following sections.

5.2.1. Electrolyser Cost

The estimation of electrolyser capital costs presents a significant challenge due to several factors, including the technology employed, economies of scale, production capacity, and future cost projections. Table 4 compiles bibliographic data on current and projected costs of alkaline electrolysis (AEL) systems, with particular focus on small-scale units around 1-2 MW. Costs are primarily dependent on the specific technology and plant capacity, with reported estimates varying up to an order of magnitude in €/kW.

The work of Reksten et al. [94] provides a useful model for estimating the capital expenditures (CAPEX) of alkaline electrolyser plants up to 10 MW, incorporating both scale factors and learning rates. Based on this model, the average current reference cost of a 1.5 MW alkaline electrolyser is assumed to be 1,000 €/kW, representing a conservative assumption for present-day installations. However, other studies [95–97] incorporate expert elicitation from both academia and industry, and account for variables such as production scalability, electricity market size, reference prices, and varying intensities of research and development. As shown by Gallardo et al. [98], using data from 2015–2017 for electrolysers ranging from kW-scale to a few MW, the cost curve is significantly steep (exceeding >9 kUS$/kW for very small systems), decreasing to approximately 0.9 kUS$/kW for scales between 1–2 MW, and exhibiting high scaling exponents up to 0.5.

Overall, the literature summarized in Table 4 indicates that current specific investment costs for MW-scale alkaline electrolysers typically range between 600 and 900 €/kW, with a median value of approximately 750 €/kW. For systems of around 1 MW, this median value can be considered representative of present market conditions. Future projections suggest further cost reductions driven by technological advancements and scaling effects, with estimates for 2030 ranging from 160 to 1,300 €/kW (median: 700 €/kW), and for 2050 range between 225 and 393 €/kW. Based on the median literature value of 750 €/kW has been adopted, resulting in an estimated electrolyser capital cost of 1.125 M€.

Table 4. Capital costs of alkaline electrolysers (AEL). Values for ≈1 MW capacity are highlighted in green.

Power / H_2 production	Electrolyser cost (€/kW)	Comments	Ref.
150 MW	600	Larger industrial demonstrations size	[99]
From small to large scale	$Actual_{2024}$: 500-1,400 $Future_{2030}$: 200-700 Used: 1,080	To feed a NH_3 plant	[55]
1 GW	730 anticipated cost in 2030	Plant in a Dutch port area by 2030	[100]
1.08 GW (19,440 kg H_2/h)	$Actual_{2018}$: 400	Bilateral Australia–Germany NH_3 project	[101]
1MW	$Actual_{2020}$: 550-1,600 $Future_{2030}$: 450-600 $Model_{2024}$: 1,000	Review of cost assessment and outlook towards 2030	[94]
1MW	$Actual_{2020}$: 1,000	Production of 3 t NH_3/h	[52]
1MW	$Actual_{2022}$: 213 – 1,300 (40% stack) $Future_{2030}$: (<1,000; 160-1,300; 700) depending estimation method	Extensive review providing cost analysis	[102]
Above 10MW	$Actual_{2020}$: 1,027; Stack 530 $Future_{2030}$: 710; Stack 357 $Future_{2050}$: 393; Stack 189	Economic modelling of different e-fuels pathways	[4]
1.2 MW	$Actual_{2021}$: 634; $Future_{2030}$: 465 Manufacturers report 2021 costs: €423	Production capacity of 160 kg H_2/d	[103]
1 MW 2.4 MW	$Actual_{2020}$: 1,000-1,200 $Actual_{2020}$: 1,000	Developments of AEL process and key variables	[104]
Multiscale MW	$Actual_{2017-18}$: 425-595 $Future_{2025-30}$: 340-425	Techno-economic analysis of H_2 AEL	[98]
1 MW	$Actual_{2020}$: 540-777	Uncertainty analysis	[95]

1 MW	Actual "Stack"2020: 237 Future "Stack"2050: < 90 Actual Investment2020: 920	Provides insights to scale up and reduce H_2 supply costs	[105]
432 kg H_2/h	Actual2020"without installation": 287	Production of 20,000 t NH_3 /y	[44]
4,167 kg H_2/h	Total investment cost: 511,161 €2019	Production rate of 3 kg NH_3/h	[106]
1-10 MW	Actual2019: 450	NH_3 synthesis	[107]
1.1-5.3 MW	Actual2018: 2,015 Most estimates: 1,100 (Range: 600-2,600)	Assesses production cost of different e-fuels	[96]
44 MW	Actual2018: 777	H_2 industry in Australia	[108]
From small to large scale	Actual2016: 400 – 1,400 Future2030: 400-1,000	Cost and performance of water electrolysis	[97]
3.2 MW	Actual2015: 925-1,630	H_2 production from solar PV	[109]

5.2.2. Energy Management Strategy and Hydrogen Storage

The energy management strategy is determined by the availability of wind energy for both the electrolyser and the remaining units of the ammonia production plant. An hourly assessment covering the period from September 2020 to March 2021— the most recent period with available wind speed data provided by the Augusto Linares buoy, located 22 nautical miles north of Santander city (43°50.67'N, 03°46.2'O)—was conducted to evaluate energy availability. The analysis indicates that, assuming a planned offshore wind farm composed of seven 1.5 MW wind turbines, the minimum hydrogen storage requirement to ensure continuous plant operation is 2,423 kg (Figure 4). The analysed period time in Figure 4 is autumn and winter time which is the part of the year with slightly higher wind velocity and, at least, 7 wind turbine are required if ammonia plant is located around Santander city. This unfavorable wind conditions shows that better locations should be searched for the ammonia plant; the seeking for this new location should be done according with the Spanish Government proposal [31] which defines the areas where offshore wind farms can be installed. Asturias and Galicia regions, which are two Cantabrian Sea communities closed to Cantabria region, have designated areas where offshore wind development is permitted. Those designated areas exhibit favorable wind conditions, enabling a reduction in the required number of wind turbines to below half of the installations proposed in the present study.

Reported values of hydrogen storage cover a wide range from 175 to 1,300 €/kg depending on the scale and the pressure [35,98,110]. For the low-pressure hydrogen storage tank chosen in the present study, a cost assumption of 350 €/kg is taken, resulting in an estimated investment of 848,050 € to ensure continuous ammonia production; this value is similar to that obtained using the cost correlation to estimate the purchase cost of a gas holder provided by Seider et al. [82].

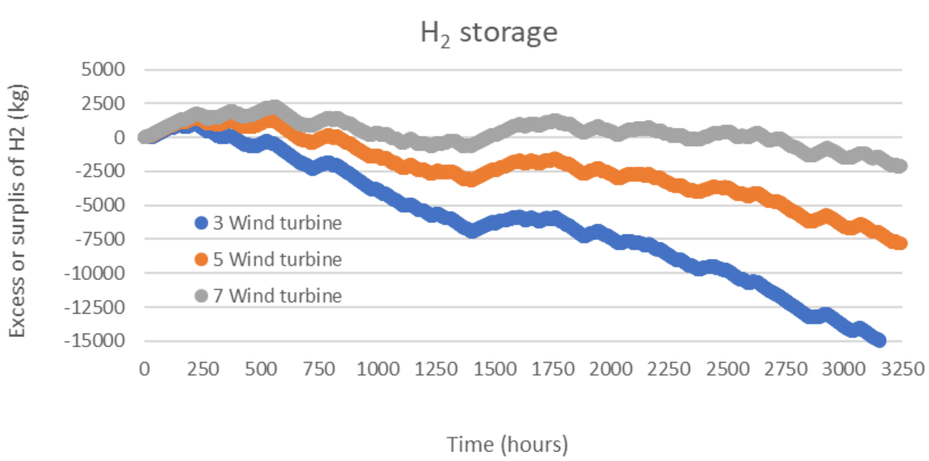

Figure 4. Hydrogen storage requirements as a function of installed offshore wind energy capacity located in the Santander coastline.

The required battery energy storage capacity to ensure uninterrupted ammonia production is four days, corresponding to the longest consecutive period with wind speeds below 4 m/s [88], since wind turbines cannot operate under such conditions. When wind speeds range between 4 and 8.5 m/s, hydrogen is produced proportionally to the available power; when wind speeds exceed 8.5 m/s, the electrolyser operates at its maximum hydrogen production capacity, enabling hydrogen storage, while surplus electrical power is used to recharge the batteries.

Additionally, the cost of energy storage for a 4-day period to maintain the electrolyser at its minimum operating capacity of 30% is estimated at 144,520 €. This estimation assumes nominal ammonia production sustained with the stored hydrogen, based on a capital cost of 100 €/kWh for lithium-ion battery racks considering high-growth projections for battery deployment [111]; a nominal capacity of 372.7 kWh [112] and a specific energy consumption of 3.55 kWh/kgNH_3, which corresponds to the fraction that must be stored to maintain 30% operation, are considered.

5.2.3. Platform Costs

The capital costs of offshore platforms are estimated based on the construction costs of offshore oil rigs. Jack-up and semisubmersible platforms are considered for fixed and floating structures, respectively. Wang et al. [37] estimate platform costs using correlations provided by Kaiser et al. [113], which are functions of production capacity, water depth, and delivery year. These correlations are defined for platforms with a typical area ranging from 4,000 to 8,500 m^2. Given a minimum draft of 9 meters at the Port of Santander and an ammonia production capacity of 2.4 tpd, these correlations yield a cost of USD 74.4 million for a jack-up platform. However, the space required for the projected plant is conservatively estimated at 200 m^2, accounting for the equipment footprint and the necessary space for piping, valves, control and safety systems, and access. This estimated area aligns with the order of magnitude of modular ammonia production units offered by companies such as AmmPower[114], FuelPositive [115], and Kapsom [116] for small-scale ammonia production. Therefore, a preliminary platform cost of 2 million € is estimated for the present study.

5.2.4. Equipment Cost Distribution

First, the Bare Module Cost of each piece of equipment was calculated as detailed in the methodology. The values obtained from these cost estimations are summarized in Table S3 provided in the Supporting Information. A total equipment cost of approximately 3.2 M€ is distributed among desalination, air separation, electrolysis, synthesis (Haber–Bosch process), and hydrogen/energy storage.

Figure 5 illustrates the Bare Module Cost (BMC) associated with the main process units involved in the green ammonia production plant proposed in this study. The cost allocation highlights the relative economic weight of each technological component in the overall system.

The electrolysis unit represents the largest investment share, accounting for 1,351,976 € (42%) of the total cost. The hydrogen storage system is the second most expensive unit, with 848,050 € (26%) due to high-pressure rating, and material constraints. This result aligns with the known capital-intensive nature of water electrolysis system to obtain H_2 and to storage it. Reducing the cost of electrolysers and H_2 storage remains one of the main challenges for improving the competitiveness of green ammonia production plants. Locating the e-NH_3 plant in an area with superior wind-energy potential, combined with anticipated decreases in electrolysis technology costs, will help reduce these expenditure categories.

The air separation unit (ASU), responsible for nitrogen production, contributes 412,733 € (13%). The ASU requires significant capital investment due to compressors and membrane separation system. The Haber–Bosch synthesis unit accounts for 355,145 € (11%) of the investment. Within this unit, the compressors constitute the largest share of the cost due to the high operating pressure required by the process. Finally, the desalination unit has the smallest economic impact, with only 105,721€ (6%) of the total cost. Although it is essential for supplying demineralized water to the

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electrolyser, its contribution is relatively low thanks to the availability and technological maturity of reverse osmosis systems. Energy storage in battery racks accounts 144,520 € and a 4% to the total BMC.

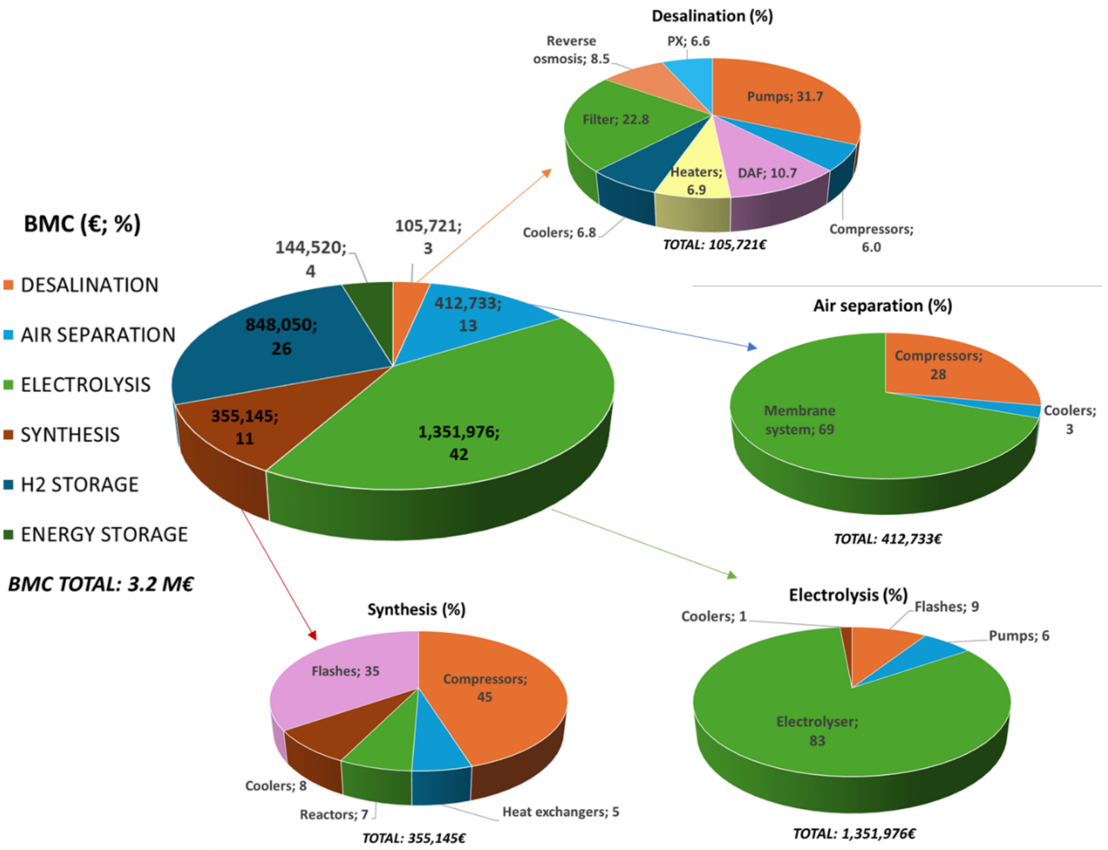

Figure 5. Equipment cost distribution among process units.

To complement the process unit distribution cost, Figure 6 shows the BMC of all the equipments classified by type. Overall, the cost breakdown confirms that hydrogen-related technologies (production + storage) dominate the equipment investment, representing more than 60% of total equipment cost. As a result, future cost-reduction strategies for green ammonia should prioritize improvements in electrolyser performance and hydrogen storage optimization.

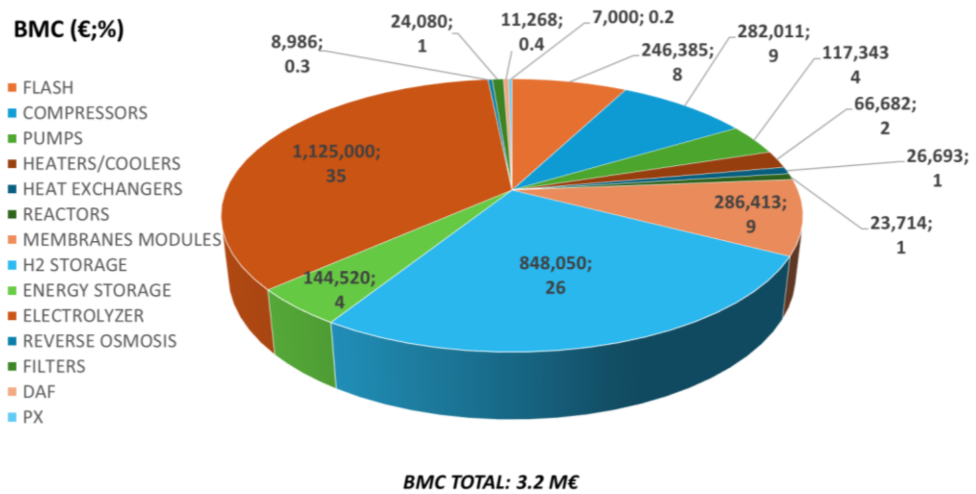

Figure 6. Equipment cost distribution by equipment type.

5.3. CAPEX and OPEX

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Once the BMC is obtained, the CAPEX can be calculated based on Eq. (1). The resulting CAPEX for the plant is 5.99 M€. When annualized, this corresponds to a value of 0.53 M€ per year. On the other hand, the resulting total OPEX is close to 1.58 M€ per year. This demonstrates that the economic performance of the process will be primarily driven by operating costs over its lifetime.

Figure 7 presents the distribution of annualized CAPEX and OPEX for the proposed plant, including the percentage distribution of each cost component. Overall, the figure highlights that OPEX represents the major contribution to the total yearly cost of the plant, significantly exceeding the CAPEX when this is annualized. This indicates that long-term operating efficiency will play a crucial role in the economic performance of the plant.

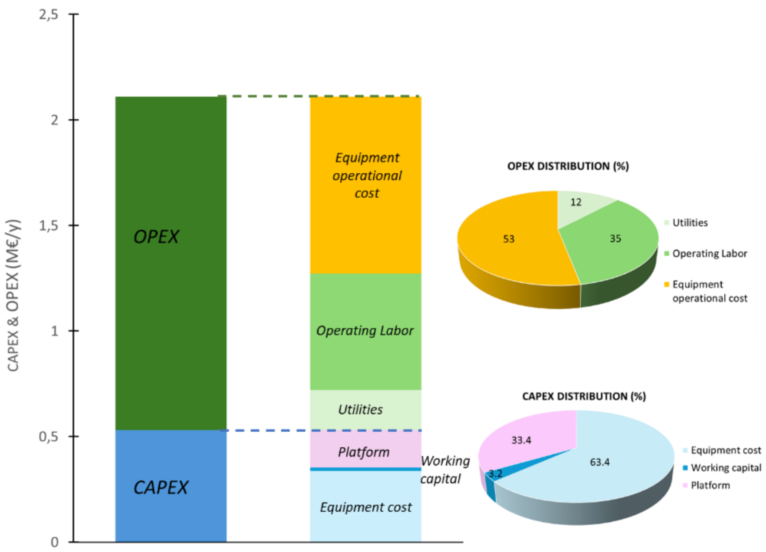

Figure 7. CAPEX and OPEX of the proposed e-NH_3 plant.

The lower section of the bars corresponds to the annualized CAPEX components, including the equipment cost investment, working capital and platform-related costs. It can be observed that the equipment investment clearly dominates the CAPEX. This result is aligned with industry expectations, as the installed equipment cost usually accounts for the largest share of the investment in chemical facilities. The platform-related costs contribute 33% of annualized CAPEX, highlighting the significant impact of site preparation.

The upper section of the bars shows the operational expenditures (OPEX). These include equipment operation costs (0.26·∑BMC), labor costs (2.18·C_{OL}) and utilities (1.075·C_{UT}). The results confirm that equipment operational costs, which can include maintenance and repair, are the most influential factor in the overall operating costs, contributing 53% of total OPEX. Nevertheless, the equation used for OPEX estimation (Eq. 3) includes multiplicative coefficients applied to the terms ∑BMC, C_{OL}, and C_{UT} that vary depending on the values associated with the individual components of these terms [83]. Therefore, a sensitivity analysis is considered necessary to assess the influence of variations in these parameters on the final cost of the plant.

Utilities represent 35% of OPEX, highlighting the energy-intensive nature of the process and the need to implement heat-integration and efficiency optimization strategies to control long-term expenses. Labor costs, although important, account for only 12%, suggesting a moderate staffing level or a high degree of process automation in a such as small-scale plant production.

Overall, the comparison between CAPEX and OPEX highlights that 75% of the total annual cost of the plant is associated with OPEX. Consequently, improvements in operational efficiency, such as optimizing maintenance, and minimizing resource and consumable use, may offer potential for cost reduction. While CAPEX-related measures can support initial economic feasibility, the long-term profitability of the plant strongly depends on controlling operating expenses throughout the lifecycle of the facility.

5.4. Levelized Cost of Ammonia (LCOA)

The levelized cost of ammonia (LCOA) is a widely used metric to evaluate the economic performance of ammonia production, as it accounts for both capital expenditures (CAPEX) and operational expenditures (OPEX) over the lifetime of the plant.

In this study, the LCOA is calculated based on CAPEX and OPEX, assuming a discount rate of 8% and a plant lifetime of 30 years. Under these assumptions, the LCOA results in 2,408.2 €/tNH_3. This value is significantly higher compared to current grey ammonia large-scale plants, which reflects the strong influence of intensive capital and operating costs associated with small-scale production, highlighting the importance of scale, financing conditions, and cost reductions for improving the economic competitiveness of e-ammonia.

In Figure 8, the obtained LCOA in the present work is compared with the values reported in the literature and previously summarized in Table 1a (offshore plants) and Table 1b (onshore plants). The LCOA value lies within the range of small-scale offshore plants (up to 10 tpd).

Most conventional ammonia plants operate at large scales, typically producing between 1,000 and 1,500 tpd, while the smallest conventional designs have capacities below 250 tpd. This tendency toward large-scale production reflects the well-known economies of scale that dominate grey ammonia synthesis for fertilizers production. A decreasing trend in LCOA values is observed as production capacity increases. This trend is slightly more pronounced in onshore plants, which indicates both economy of scale and level of technology maturity.

The green band in Figure 8 represents large-scale conventional grey ammonia plants, covering a production range of approximately 600 to 3,300 tpd NH_3. For this conventional large-scale ammonia production based on natural gas reforming, reported LCOA values range from 260 to 500 €/tNH_3, depending mainly on natural gas prices, electricity costs, and plant configuration [34–36]. These values effectively define the current benchmark for mature, fossil-based ammonia production.

Several studies estimating the production, storage, and distribution of e-NH_3 by 2050, the target year set by the European Union to achieve climate neutrality, indicate that LCOA values for large-scale plants are expected to range between 200 and 610 €/tNH_3 [105,117–120], showed in the textured band in Figure 8. Electricity costs are identified as a critical factor in reducing these values compared to current levels, together with technological innovations and economies of scale. Nevertheless, these authors emphasize that the economic viability of e-NH_3 requires an integrated approach that, in addition to technological advancements, includes fiscal incentives and a supportive regulatory framework for its deployment. The broader range for 2050 reflects the possible variability in costs (e.g., electricity prices, electrolyser capital costs). The lower bound of this range suggests that, under favorable renewable electricity prices and significant technological maturity of hydrogen technology, green ammonia could become cost-competitive with conventional production in the long term.

It is expected that small-scale facilities will also benefit from reductions in unit costs, given the growing importance of distributed NH_3 production in modular flexible plants designed to supply maritime fuels and capable of accommodating the intermittent output of nearby renewable-energy installations.

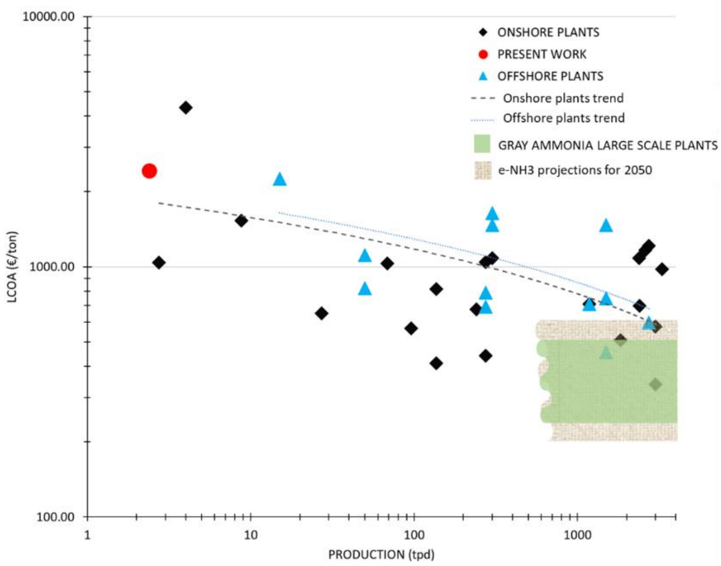

Figure 8. LCOA versus e-NH_3 plant production based on literature data included in Tables 1a and 1b.

5.5. Sensitivity Analysis

The sensitivity analysis is carried out under two approaches. First, the inherent variability in estimates derived from modular costing methodologies—universally applied to preliminary process designs—can introduce variations of up to 50% [79] in CAPEX and OPEX estimations. Second, there are a series of individual variables with a hierarchical impact on costs that should be considered in a dynamic and globalized market.

Firstly, the sensitivity analysis was carried out by varying major capital and operating expenditure parameters by ±50% with respect to the base case. The objective of this analysis is to assess the relative influence of each cost category on the levelized cost of ammonia (LCOA) and to identify the dominant economic drivers of the system. The results are summarized in Figure 9, where the LCOA response is plotted as a function of the percentage variation in the main CAPEX and OPEX components. Results indicate that both OPEX and CAPEX have a strong influence on the LCOA, varying in the range of 1,500-3,300 €/tNH_3.

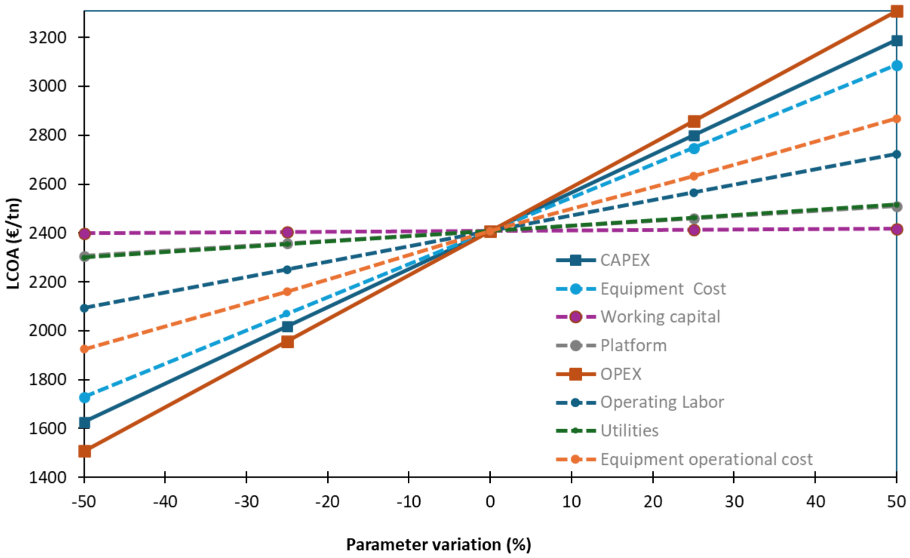

Figure 9. Sensitivity analysis for variation of CAPEX and OPEX (Baseline = 2,408.2 €/tNH_3).

Among the cost components, the equipment cost is the most sensitive parameter affecting the LCOA, as it shows the sharpest slope among all evaluated cost components. This highlights the capital-intensive nature of the plant and the importance of equipment selection, scaling strategies,

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and potential cost reductions. Also, the equipment operation costs exhibit a strong sensitivity, more than that of utilities and operating labor. This is explained by the fact that equipment operational costs are assumed to be 26% of the total equipment investment cost, directly linking this OPEX component to the CAPEX component.

In contrast, utilities and operating labor costs show a low impact on the LCOA. In particular, the utilities curve almost overlaps with that of the platform cost. Working capital also has a very limited effect on the LCOA, as its value remains nearly constant over the analyzed range. Consequently, its influence on the overall economic performance of the system can be considered negligible. Overall, the sensitivity analysis demonstrates that the LCOA is primarily driven by capital-intensive components, with equipment cost being the most critical parameter.

Together with CAPEX and OPEX sensitivity analysis, a second approach to the sensitivity analysis based on specific economic parameters was carried out represented through a tornado diagram. The selected parameters were electrolyser price, electricity cost, operator salary, hydrogen storage cost, discount rate, platform cost and plant lifetime, as the variables expected to have the greatest influence on LCOA and the highest uncertainty in their estimation (Figure 10).

As it was discussed before, OPEX and CAPEX (±25%) are the most influential parameters, producing the largest variation in LCOA. The electrolyser cost was varied between 160 and 1300 €/kW, according to the projections for electrolyser technologies by 2030. This wide range reflects uncertainty associated with technology and it significantly impacts the overall ammonia production cost, varying the LCOA almost ±400 €/tNH_3.

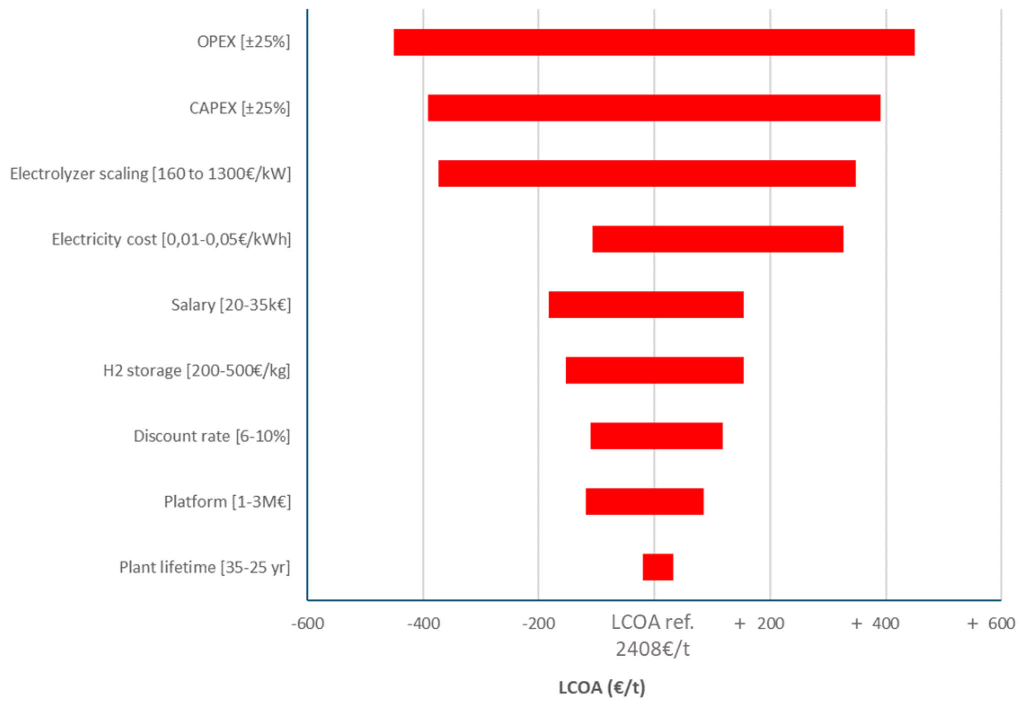

Figure 10. Effect of variation in selected economic parameters on the LCOA.

The electricity price was varied between 0.01 and 0.05 €/kWh, representing values commonly reported in the literature for renewable electricity. The increase in electricity costs has a significant impact on the LCOA, leading to increases of approximately 100 €/tNH_3 for each 0.01 €/kWh rise in electricity price. The potential use of grid electricity (0.06 €/kWh) would result in an increase of about 400 €/tNH_3.

Operating labor costs were assessed by varying the average operator salary between 20,000 and 35,000 € per year [121], reflecting typical ranges reported for industrial plant operators in Spain. The effect of this parameter on the LCOA was around ±180 €/tNH_3.

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Additional parameters, including hydrogen storage cost (200–500 €/kg), discount rate (6–10%), platform investment (1–3 M€), and plant lifetime (25–35 years), show progressively lower influence on the value of LCOA.

6. Conclusions

Electro-ammonia production plants are gaining increasing attention as key enablers of the energy transition. The integration of ammonia synthesis with renewable energy sources offers a pathway to decarbonize both the chemical industry and the energy sectors. In this context, decentralized and flexible small-scale production plants offer sustainable solutions tailored to specific contexts, allowing production to be located close to renewable energy generation sites and near end users. However, the competitiveness of small-scale plants is largely limited by economic factors and, therefore, techno-economic analyses of small-scale green ammonia plants are essential to identify the key parameters for viability and to guide future design strategies aimed at improving competitiveness.

This work presented a preliminary techno-economic assessment of a small-scale e-ammonia production plant with a capacity of 2.4 tpd (100 kg/h) located at the Port of Santander, Cantabria region, Spain. It should be mentioned that the area is characterized by unfavorable wind conditions, which increase storage costs and constrain the future expansion of wind-energy generation. The results confirmed that, under current technological and economic conditions, small-scale green ammonia plants are not yet competitive compared to conventional grey ammonia production plants, either in terms of capital expenditure (CAPEX) or operating expenditure (OPEX). The estimated CAPEX of 5.99 M€ (equivalent to 0.53 M€/y), OPEX of 1.58 M€/y, and a Levelized Cost of Ammonia (LCOA) of 2,408 €/tNH_3 are higher than most of the values reported for ammonia plants in the literature. These higher values are mainly driven by the lack of economy of scale and the high cost of green hydrogen production and storage. Nevertheless, the results of this study provide valuable information for decision-making related to the implementation of flexible, modular e-ammonia production plants operating under mild conditions. Therefore, this contribution will support progress in industrial competitiveness and in meeting societal decarbonization targets.

Despite the current economic disadvantage, green ammonia production plants are expected to become more competitive in the future. Literature consistently highlights that improvements in electrolyser efficiency, reductions in renewable electricity costs, and the advancement of modular and standardized plant designs, will play a decisive role in lowering production costs.

A key aspect is the availability of renewable energy. The data obtained highlight the importance of superregional techno-economic assessments that consider the optimal availability of offshore wind energy within an interregional collaborative framework involving all relevant stakeholders. The development of interregional innovative ecosystems and environments, engaging renewable-energy suppliers, raw-material and process-technology providers, R&D centers, financial institutions, certification bodies, and companies involved in the transport, storage, and use of e-fuels, forms the basis of forthcoming European and national R&D funding calls in which further work will be required.

In this transition context, small-scale plants have particular strategic importance. Although they do not fully benefit from scale economies, they offer advantages in terms of energy integration. Therefore, continued research and development, along with supportive policy frameworks, will be critical to enable the adoption of green ammonia as a sustainable energy carrier and chemical feedstock.

In addition to the techno-economic optimization of the process, it will be necessary to evaluate its sustainability through a life-cycle assessment and to determine the potential contribution of such plants to achieving the Sustainable Development Goals (SDGs).

Supplementary Materials: The following supporting information can be downloaded at: https://www.mdpi.com/article/doi/s1, Table S1. Methodologies applied for Cost estimation of equipment; Table S.2. Dimension parameters of equipment; Table S.3. Bare Module cost of equipment.

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Abbreviations

The following abbreviations are used in this manuscript:

ACM	Aspen Custom Modeler
AEL	Alkaline electrolysis
ASU	Air separation unit
BMC	Bare Module Cost (€)
CAPEX	Capital Expenditure (€)
CEPCI	Chemical Engineering Plant Cost Index
C_{OL}	Cost of operating labor (€/y)
$C_{platform}$	Cost of platform (€)
C_{RM}	Cost of raw materials (€/y)
C_{UT}	Cost of utilities (€/y)
C_{WT}	Cost of waste treatment (€/y)
DAF	Dissolved Air Flotation
DMC	Direct Manufacturing cost (€/y)
DMF	Dual Media Filtration
e-NH_3	electro-ammonia
F	yearly ammonia production (t/year)
FMC	Fixed Manufacturing costs (€/y)
GE	General Expenses (€/y)
GHG	Greenhouse Gas
HB	Haber–Bosch
HHV	Higher heating value (MJ/kg)
HX	Shell-and-tube heat exchanger
i	Discount rate (%)
IEA	International Energy Agency's
ICVE	Internal Combustion Engines
IMO	International Maritime Organization
LCOA	Levelized cost of e-NH_3 (€/tNH_3)
MOD	Membrane module
Nnp	Number of non-particulate processing steps
OPEX	Operational expenditure (€/y)
P	Number of processing steps involving solids handling
PL	Plant lifetime (years)

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PtF	Power-to-Fuel processes
PtX	Power-to-X (PtX) processes
PX	Pressure exchanger
RO	Reverse osmosis unit
RX	Reaction stage
SDG	Sustainable Development Goal
SOFC	Solid oxide fuel cell
TDS	Total dissolved solids (ppm)
tpd	tons per day
TPI	Total Permanent Investment (€)
UF	Update Factor
WC	Working Capital (€)

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