Techno-economic analysis of green hydrogen production - 20 MW PEM electrolysis plant

"Hydrogen economy in the food system" project Report on Task 1.3 & 1.4

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1 Introduction

Hydrogen is expected to play a crucial role in deep decarbonization strategies across various sectors. To align with climate goals, hydrogen must be low-carbon from the outset and ultimately produced using renewable energy sources—commonly referred to as green hydrogen (IRENA, 2020). Among the available production methods, water electrolysis powered by renewable electricity is considered the most sustainable pathway. However, a major barrier to the widespread adoption of green hydrogen is its relatively high production cost compared to hydrogen derived from fossil fuels (Ibáñez-Rioja, 2023). The single largest cost driver in on-site green hydrogen production is the price of renewable electricity required to operate the electrolyzer.

Wind power harnesses kinetic energy from the wind and converts it into electrical energy (Zhao et al., 2018), offering a plentiful and renewable energy source. In Finland, wind power development has accelerated significantly since 2012. By the end of 2024, the country had installed 1,835 wind turbines with a total capacity of 8,358 megawatts (Renewables Finland, 2025), supplying approximately 24% of Finland's electricity consumption. Looking ahead, wind power production is projected to reach 100 TWh from onshore and an additional 100 TWh from offshore sources by 2040, underscoring Finland's significant potential for continued expansion in renewable energy generation.

Wind power development is also advancing rapidly in South Ostrobothnia, where up to 4,400 MW of new capacity is planned by 2030 (Spoof-Tuomi, 2024). This expansion presents a significant opportunity for the region to harness abundant and cost-effective, locally produced renewable energy for green hydrogen production via water electrolysis.

This paper investigates the technical concept and hydrogen production costs of an electrolysis plant integrated with wind power. The report is structured into two main sections. The first section presents an engineering model of a selected commercial-scale hydrogen production process based on water electrolysis, coupled with a wind power source. Initially, a theoretical electrolyzer model was developed using established literature. This model includes not only the electrochemical stack but also all relevant balance-of-plant components and auxiliary systems. To assess its accuracy, the theoretical model's performance was compared with the results from a simulation model developed using MATLAB/Simulink. The second section focuses on the economic modelling of the selected electrolyzer-based hydrogen production system. This analysis is conducted using the Levelized Cost of Hydrogen (LCOH) methodology. Three operational scenarios are considered: (1) the electrolyzer operates solely on electricity from a dedicated wind farm, (2) the electrolyzer utilizes only surplus electricity from the wind farm, and (3) the electrolyzer is connected to the grid and uses only grid electricity. To provide a more comprehensive understanding of the economic outcomes, sensitivity analyses were performed on key variables.

The structure of this paper is as follows. Chapter 2 outlines the rationale for selecting electrolysis technology and introduces the fundamental principles of proton exchange membrane (PEM) electrolysis. It also defines the facility's location and production capacity. Chapter 3 provides a more detailed description of the PEM electrolysis process, including all auxiliary systems and components required for its proper operation. Additionally, this chapter presents cost data for the individual components. Chapter 4 describes the simulation model developed using MATLAB/Simulink, along with the corresponding simulation results. Chapter 5 focuses on the economic analysis of the system, while Chapter 6 summarizes the key findings and presents the final conclusions of the study.

2 System overview

2.1 Rationale behind the electrolysis technology choice

The most suitable electrolysis technology to be coupled with renewable energy depends on various factors, including the specific requirements of the application, economic considerations, and technical feasibility. Currently, two leading electrolysis technologies used for hydrogen production are alkaline electrolysis and proton exchange membrane (PEM) electrolysis.

Alkaline electrolysis systems are known for their maturity, robustness, long lifespan, and low capital costs compared to other electrolyzer technologies. However, these conventional systems are not designed to be operated by combining fluctuating power sources (Kojima et al., 2023). For example, alkaline electrolyzers suffer from slow system response times to load changes (Yodwong et al., 2020), which could be a consideration when coupling with intermittent renewable energy. A significant weakness of alkaline electrolysis is that the minimum load of the alkali electrolysis is typically limited to 15-25 % of the nominal hydrogen production due to the lateral diffusion of hydrogen through the diaphragm to the oxygen side, resulting in the formation of a combustible mixture at low production. In practice, safety stops occur with 1-2 % hydrogen contamination in the oxygen stream (Buttler & Spliethoff, 2018). Start and stop operations result in severe catalyst degradation (Kojima et al., 2023). Another weakness of alkaline technology is its relatively low current densities (0.2-0.4 A/cm²), leading to low performance profiles (Anwar et al., 2021) and making these electrolyzers large and heavy compared to other technologies for similar production rates (de Groot et al., 2022).

On the contrary, proton exchange membrane electrolysis, also a commercial electrolysis technology, is known for its fast response times, making PEM electrolysis a well-suited technology for dynamic operation associated with intermittent energy sources. Moreover, the very low gaseous permeability of the polymeric membranes lowers the risk of the formation of flammable mixtures; hence, operating at the full load range (0–100 %) is possible (Ursua et al., 2012). Further advantages of PEM electrolysis include high current density due to the low ionic resistance of polymer membranes, compact module design due to the solid electrolyte, small footprint, and fast start-up. (Buttler & Spliethoff, 2018; Kumar & Himabindu, 2019). It also shows less performance degradation with switch-on/off actions than alkali electrolysis (Kojima et al., 2023). The main challenge of PEM electrolyzers is the high investment cost, which can be a disadvantage for large-scale hydrogen production.

The choice between these most advanced and already commercial electrolysis technology options depends on several project-specific considerations, such as response time, scale, and cost. This study prioritized short response time and minimum operational restrictions. As a result, PEM electrolysis was identified as the most suitable option and selected as the preferred technology.

2.2 Principle of PEM water electrolysis

The core of the electrolyzer is electrolysis cells. This is where the main electrochemical process takes place. The fundamental design of a PEM electrolysis cell is shown in Fig. 1. The heart of the cell is the membrane electrode assembly (MEA) that separates the cell into two half-cells. MEA combines the solid polymer electrolyte membrane and the two electrodes — a positively charged anode and a negatively charged cathode. The PEM itself is a thin, selective membrane typically made of perfluorosulfonic acid (PFSA) polymer. PEM is responsible for transporting ions from one electrode to the other and, at the same time, physically separating the produced gases (IRENA, 2020). Catalysts are used to increase the current density and the rate of electrolysis reactions (Dincer &

Acar, 2015). Other key components are porous gas diffusion layers (GDL) and bipolar plates on both sides of the cell. GDLs facilitate water and gas transport, as well as thermal and electrical conduction between the electrodes and bipolar plates (Kang et al., 2020). Bipolar plates provide multiple functions, such as conducting electrons, supporting the membrane and electrodes, connecting single cells to realize a stack, and introducing thermal conductivity to handle the PEM temperature (Yodwong et al., 2020).

In PEM electrolysis, deionized water is fed to the anode side of the cell, distributed by GDL and reaches the catalyst layer. When direct current (DC) electricity is applied between two electrodes, water is dissociated into (H⁺), electrons (e⁻), and oxygen (O₂) at the anode. The O₂ formed during this half-cell reaction is removed with the unconsumed water, while protons move through the proton-conducting membrane towards the cathode side. The electrons exit from the anode through the external power circuit, which provides the driving force (cell voltage) for the reaction. At the cathode side, the protons and electrons recombine to form hydrogen gas. (Kumar & Himabindu, 2019; Sood et al., 2020)



Anode: $2H_2O \rightarrow O_2 + 4H^+ + 4e^-$ Cathode: $4H^+ + 4e^- \rightarrow 2H_2$ Overall: $2H_2O(l) \rightarrow 2H_2(g) + O_2(g)$

Figure 1. Fundamental design of a PEM electrolysis cell (Toghyani, 2019)

2.3 Balance of plant

In addition to the cells/stacks, an electrolysis system includes various auxiliary systems and components necessary for the proper function of the electrolysis process. The typical components of a PEM system are shown in Figure 2 and typically include (Tsotridis & Pilenga, 2018):

- > Power supply, consisting of
 - grid connection and transformer to adjust the electricity to the operational requirements
 - rectifier to convert alternating current (AC) into direct current (DC)
 - system control board (safety sensors, process parameter measuring devices, etc.)
- > Water conditioning for the treatment of the water supplied and recovered
 - make-up water tank
 - feed water pump
 - deionised water (DIW) production unit
 - anodic circulation loop consisting of an ion-exchanger to keep the water quality at the desired level, an oxygen/water separator vessel, and a demister for further removal of small liquid-water droplets from the gas outlet stream

- cathodic circulation loop consisting of a hydrogen/water separator vessel and the subsequent demister
- Process utilities such as the water recirculation pump and process-value-measuring devices (pressure sensor, flow meter, gas sensors)
- Process cooling, consisting of heat exchangers for the thermal management of the process water (to remove heat from the circulation loop) and to keep the stack at the proper temperature range
- Gas cooling, consisting of heat exchangers for the thermal management of the gases produced during the electrolysis process
- Gas purification to clean the hydrogen product stream to the desired level of quality. Gas purification step typically consists of
 - a deoxidizer to remove traces of oxygen that could be present due to crossover effects
 - a gas dryer to remove residual moisture
 - a buffer tank for compensation of variable hydrogen production
- > Gas compression and storage, composed of
 - pressure control valves for hydrogen and oxygen to operate the electrolyser system at the desired pressure level
 - a compressor to bring the hydrogen gas pressure to the specified value
 - high-pressure storage tanks for the final storage of the hydrogen produced by the electrolyser



Figure 2. Schematic representation of the components of a PEM water electrolysis system (Tsotridis & Pilenga, 2018)

These components also consume energy and directly affect the electrolyser's performance and efficiency (Sood et al., 2020). Please note that in Figure 2, the water feed to the cathode side is optional, as it is only there to facilitate the efficient removal of the hydrogen.

2.4 Electrolysis plant siting

When evaluating the optimal location of the electrolysis plant, the four primary considerations were:

- 1. Proximity to renewable energy sources
 - Locating the electrolysis plant close to a reliable and abundant renewable energy source is crucial to reduce transmission losses and ensure a cost-effective operation.
- 2. Availability of water resources

Water is a central input for the process, so access to a stable and sufficient water supply is essential. Potential water sources are surface water like rivers and lakes, groundwater, seawater, water from the public grid, and recycled water, e.g., treated industrial wastewater (Simoes et al., 2021). However, it is important to note that some of these sources may also be used for agriculture and human consumption, and the use of these sources for hydrogen production should not compromise their availability for other essential purposes.

- 3. Market opportunities and proximity of potential customers Potential customers include industry, transportation, power generation, etc.
- 4. Infrastructure and logistics

A location with well-established infrastructure, including transport networks, roads and utilities, can reduce construction and operational costs.

In South Ostrobothnia, up to 927 megawatts of new wind power capacity are planned in the vicinity of Hirvijärvi artificial lake between the cities of Seinäjoki and Lapua by 2026–2027 (Finnish Wind Power Association, 2024). This could be one possible location for an electrolysis plant that meets the above criteria.

2.5 Plant capacity

The optimal size of an electrolysis plant is a complex decision that involves balancing technical, economic, and market-related considerations. In this case study, for the sake of simplicity, it was assumed that about 10 % (100 MW) of the new wind power capacity would be connected to hydrogen production. To determine the optimal electrolysis plant capacity, we first forecasted the wind power generation based on historical wind data for the selected location. Hourly data from 2023 were retrieved from Renewables.ninja (2024), a web tool developed by Imperial College London and ETH Zürich that displays local wind speed data and the estimated amount of energy that wind turbines could produce. A hub height of 150 m and blade diameter of 136 m were used for wind turbines. The hourly values were then averaged weekly. Figure 3 shows the predicted wind power generation profile scaled to match the nominal wind power capacity of 100 MW. The minimum wind speed required for most modern wind turbines to start turning and generate electricity, the so-called cut-in speed, is typically around 3 to 4 m/s. Thus, power generation with wind speeds below 3.5 m/s was set to zero, giving the annual operating hours of 8426 h for wind turbines. The total capacity factor was 42.3 %.



Figure 3. Wind power generation of a 100 MW wind farm; weekly average based on 2023 wind data

The main goal in electrolysis plant sizing was a high electrolyser load factor. A high electrolyser load factor increases the quantity of hydrogen produced and helps reduce hydrogen production costs (Nock et al., 2021). Figure 3 shows that sizing the electrolyser capacity to 20% of the wind power capacity results, in principle, in an electrolyser load factor of 100%. Therefore, the electrolyser capacity was set at 20 MW.

3 Electrolysis plant layout and component specifications

The PEM system modelled in this paper is intended to represent an average large-scale PEM that would be built today. The plant layout follows the model described by Holst et al. (2021). Figure 4 shows a general schematic of a PEM system.



Figure 4. Plant layout of a PEM electrolysis system (Holst et al., 2021)

The heart of the plant is the PEM electrolysis stacks. DC power is supplied to the stack by a rectifier, which converts AC to DC. The water is fed into the system from the water purification system. A feed water pump increases the water pressure to the pressure of the anode side and feeds it to the process. On the anode side, the oxygen formed during the electrolysis process is separated from the unconsumed water in the gas/water separation vessel. The oxygen leaves the system through a demister, which removes fine water droplets from the oxygen stream, and the unconsumed water is recirculated in the process. A heat exchanger in the anode circulation loop removes process heat from the system to prevent damage to the stacks. The heavy metal ions that dissolve from the stack components and the system during operation are removed in the ion exchanger. There is no separate water circulation system on the cathode side. However, due to a drag current, water gets from the anode side to the cathode side. For this reason, a smaller gas-water separator is also installed on the cathode side. The water that has accumulated in the gas-water separator is returned to the anode side via a drain valve. The hydrogen gas stream is led through a heat exchanger, where a large part of the water entrained condenses and is separated via a condensate separator. To use the hydrogen for subsequent applications or storage, it is dried and the oxygen entrained is removed. A deoxidizer reactor performs oxygen removal, and the final fine drying is done through pressure swing (PSA) or temperature swing adsorption (TSA).

3.1 Schematic layout and the key parameters of the system

The schematic layout of the investigated PEM electrolysis plant is shown in Figure 5. The 20 MW PEM electrolyzer is divided into two stack arrays. Both stack arrays consist of 10 PEM electrolysis stacks with a capacity of 1 MW. Each electrolysis stack is connected to an individual rectifier. The plant has a central water purification unit, which supplies both stack arrays. Also, both arrays are connected to a central gas purification subsystem and a central cooling subsystem.



Figure 5. Schematic layout of 20 MW PEM electrolysis plant. Modified from Holst et al. (2021)

The key PEM parameter values applied in this paper are summarized in Table 1. According to Faraday's law, the hydrogen production rate is proportional to the current density. Nevertheless, for a real electrolysis cell, the Faraday efficiency η_F (the ratio of the actual to theoretical H₂ production rate) must be introduced. The difference between actual and theoretical production results from parasitic current losses in gas pipes and cross permeation of gaseous products. So, the maximum H₂ production capacity was calculated as follows:

$$\frac{dm}{dt} = \eta_F \frac{j * A * M}{z * F}$$

where dm/dt is the hydrogen mass flow in g/s, η_F is the Faraday efficiency (99 %), *j* is the current density in A/cm², A is the cell area in cm², *M* is the molar mass of hydrogen (2.016 g/mol), *z* is the charge number of ions (for hydrogen *z*=2), and *F* is the Faraday constant 9.648533 x 10⁴ As/mol.

A voltage degradation factor was also included to account for the electrolyzer performance losses over time. The main degradation phenomena for PEM water electrolyzer components include structural changes to the catalysts, deactivation of the electrolyte, loss of performance due to ion impurities in the feed water, and the corrosion and passivation of titanium-based bipolar plates and gas diffusion layers (Liu et al., 2023). According to the JRC report (Davies et al., 2021) several new projects appear to have achieved the 2030 degradation target of 0.12 %/1000 h. The voltage degradation factor of 0.12 %/1000 h, corresponding to 2 μ V/h, was also used in this study.

Parameter	Value	Unit	Description
Plant capacity	20	MW	
Stack per system	20		
Stack unit capacity	1	MW	Capacity of single-stack
Min. load	5	%	Operational limitation
Max. load	100	%	Operational limitation
Voltage efficiency (% of LHV) Voltage efficiency (% of HHV)	65.2 77.2	% %	
Specific energy consumption	51.05	kWh/kg of H ₂	Specific energy demand of electrolysis stacks at rated load (100%) at start-of-life
Max. H ₂ production capacity	393	kg/h	
Plant availability	96	%	
Purity (with de-oxo dryer)	99.9995	%	
Pressure cathode	30	bar	The output pressure of hydrogen
Pressure anode	1	bar	The output pressure of oxygen
Temperature at nominal load	70	°C	
Feed water consumption	10	kg H_2O/kg of H_2	Quantity of deionized water required to produce 1 kg of hydrogen
Voltage degradation	2	μV/h	Electrolyser performance losses over time

Table 1. Key parameters applied in this study

The following subsections present more detailed descriptions of the process components. The related investment costs and the energy needs of different processes are also determined for techno-economic modelling.

3.2 Electrolyser cells/stacks

Electrolysis stacks follow a modular structure. A stack consists of several cells connected electrically in series and hydraulically in parallel, with the bipolar plates separating adjacent cells (Figure 6). Due to the high pressures inside the stack, massive end plates made of steel are applied to ensure gas tightness. Each cell has a power rating (kW) and these cumulative cells equal the stack's total capacity. The stack specifications presented in Table 2 were chosen within the range of typical values in the literature.



Figure 6. Schematic design of PEM electrolyser cell and stack. Modified from Ruth et al. (2017) and (Holst et al., 2021)

Current density and reference voltage at the cell level are critical parameters determining cell efficiency and energy consumption (Mayyas et al., 2019). In this study, the current density of 2.0 A/cm² and cell voltage of 1.9 were used, based on Holst et al. (2021). The single cell active surface area was set at 1000 cm², also based on Holst et al. (2021). The power density, determined from current density and cell voltage, was then used to calculate the number of cells per stack to match it with the stack size (1 MW). Other stack component specifications were collected from Mayyas et al. (2019) and Holst et al. (2021).

Specification	Value	Unit
Stack size	1	MW
Current density	2.0	A/cm ²
Reference voltage	1.9	V
Active surface area per cell	1000	cm ²
Power density	3.8	W/cm ²
Single-cell power	3800	W
Faradaic efficiency	99	%
Cells per stack	264	
Cells per system	5280	
Stack lifetime	80 000	h
Membrane (Nafion 117) thickness	178	μm
Catalyst loadings Anode: Platinum Cathode: Iridium oxide 	1.0 2.0	mg/cm² ma/cm²
 GDL, anode side sintered porous titanium (Ti) gold (Au) coating GDL, cathode side Carbon cloth 	1.0 100 280	mm nm µm
 Bipolar plates (BPP) Structured titanium plates Tantalum (Ta) coating (anode side) 	1.0 1.0	mm µm
End plates (pressure plates) Stainless steel 1.4301 / AISI 304	20	mm

Table 2. Electrolyser cell and stack specifications

The electrolyzer stack cost model followed the method introduced by Krishnan et al. (2023). Accordingly, the stack cost was divided into the following categories:

- 1. Direct material cost
- 2. Direct manufacturing cost
- 3. Manufacturing labour cost
- 4. Other direct manufacturing cost, including capital and interest, energy cost, and equipment maintenance cost
- 5. Indirect overhead cost, including R&D, prototype production and engineering, sales and marketing, and administrative expenses

The most dominant contribution to the direct cost is coming from material cost (190 €/kW) (Krishnan et al., 2023); see Figure 7. The stack components that contribute most to the material cost are the membrane-electrode assembly (MEA) and gas diffusion layers (Mayyas et al., 2021). Regarding other

direct costs (labour and manufacturing cost) and overheads, the costs also followed the values presented by Krishnan et al. (2023), which are based on financial reports from electrolyser manufacturers. Thus, a total specific capital cost of 617 \in /kW was applied for a 1 MW PEM stack. The total cost of 20 stacks was 12,340,000 \in .



Figure 7. Cost breakdown for a 1 MW PEM electrolyzer stack. Modified from Krishnan et al. (2023) and Mayyas et al. (2019)

3.3 Stack power supply subsystem

Whether the electricity source is the grid or on-site wind power generation, transformers, AC-DC converters (rectifiers) and other power equipment will be required to meet the electrolyzer's electricity demand. Rectifiers are needed to convert the alternating current (AC) to direct current (DC) and to control the power quality. In addition, current and voltage sensors are needed to continuously monitor the electrical supply to the stack. Figure 8 shows a conceptual schematic of a wind-hydrogen plant.



Figure 8. Conceptual schematic of the wind-hydrogen plant (Olateju et al., 2014)

The power supply system plays a key role in determining the overall efficiency and performance of the electrolysis system. PEM electrolysis requires precise voltage and current control to optimize the electrochemical reactions occurring at the anode and cathode. In PEM electrolysis systems, the most suitable rectifier is typically a high-quality, dedicated DC power supply that can provide stable and controlled direct current to the electrolysis cell.

In principle, two main topologies are possible for rectifiers: thyristor-based solutions and insulatedgate bipolar transistor (IGBT) power electronics (ISPT, 2022). Currently, the most commonly used topology in industrial-scale electrolysis systems is the thyristor-based topology. However, this topology presents certain challenges in terms of power quality, leading to higher stack-specific energy consumption and additional losses (Keddar et al., 2021). Simulation results have shown that transistor-based converter topology can offer up to 9–14 % lower stack-specific energy consumption than the traditional 6- or 12-pulse thyristor rectifiers (Koponen et al., 2019). Therefore, rather than traditional thyristor-based topology, the IGBT topology was applied in this study. Although IBGT is more expensive, it is expected to be cost-competitive with thyristor-based solutions at the system level because, unlike thyristors, IGBT technology uses active control and does not require harmonic filters and power-control devices to provide reactive power compensation (ISPT, 2022).

In the PEM model created in this study, each electrolysis stack is connected to an individual rectifier. The output voltage window of rectifiers must include the minimum voltage of the electrolysis stack at the start-of-life and partial load, as well as the maximum voltage of the stack at the end-of-life at full load (Holst et al., 2021). This is because the degradation of electrolysis cells leads to higher cell voltage with increasing operation hours (the stack voltage must be increased to produce the rated amount of hydrogen even with degraded stacks). Assuming a cell voltage of 1.9 V, a linear voltage degradation of 2 μ V/h over a lifetime of 80,000 h results in a cell voltage of 2.06 V at end-of-life.

In addition to rectifiers, power electronics include safety sensors and various process parameter measuring devices. For the balance of plant, power supply represents a significant cost component, representing 15–20 % of the capital cost (CAPEX) of electrolyser systems (Keddar et al., 2021; Koponen et al., 2019). Capital cost estimates found in the literature for power supply and electronics are presented in Table 3. All capital cost estimates are indexed to EUR₂₀₂₃ currency.

Capital cost for power supply and electronics	Stack size	Reference
222 € ₂₀₂₃ /kW	1 MW	Holst et al., 2021
255 € ₂₀₂₃ /kW	1 MW	IRENA, 2020
204 € ₂₀₂₃ /kW	1 MW	Mayyas et al., 2019

Table 3. Capital cost estimates for power supply and electronics

To comply with the cost estimates presented above, this study uses a capex cost of 227 €/kW for power supply and electronics. Thus, the stack-specific cost of power supply is 227,000 €, and the total cost for 20 stacks is 4,540,000 €.

3.4 Water collection, treatment and storage

Total water requirements depend on several factors, including water quality and the treatment method used. From a stoichiometric point of view, the H_2 production via water electrolysis consumes 9 kg of water per 1 kg of H_2 . However, most commercial electrolysis units on the market today advertise that they require between 10 and 11 kg of deionised water per kg of hydrogen produced (Saulnier et al., 2020). In addition, water supply factors such as losses during water collection and treatment processes must be considered (Simoes et al., 2021).

The water source in this study is surface water from the nearby lake. The water collection, treatment, and storage involve the following steps:

- Raw water collection and pumping via a pipeline to the hydrogen production plant
- Raw water treatment, including fine screening to remove solid materials, coagulation and filtration to remove particulates and turbidity, and reverse osmosis to remove salts (ions), colloids, organics, and bacteria from the water.
- Disposal of residues from water treatment
- Storage of the treated water

Each of these steps involve energy consumption and costs, as well as water losses. The energy consumption of water collection and treatment is based on the following inputs and assumptions:

- The sum of total raw water needs for the electrolyser was set 65 % on top of the stoichiometric water demand of the electrolysis reaction. This value includes 10 % water losses in water collection and transport, 30 % losses in water treatment processes, and 10 % water losses due to evaporation and leaks. Moreover, an additional 15 % was assumed to be needed for minimising shortage risks. Thus, the total raw water demand was estimated at 15 litres water per kg of H₂ (= 6 m³ of raw water per hour)
- The power requirement *P* (in watts) for water transport (pumping) was calculated as follows:

$$P = \frac{\rho * g * Q * H}{\eta_M * \eta_P}$$

where ρ is the fluid density in kg/m³, *g* is the gravitational acceleration (9.81 m/s²), *Q* is the fluid volumetric flow rate in m³/s, *H* is total dynamic head (energy required to overcome friction losses, elevation changes, and other hydraulic resistances in the system) in meters, η_M is the efficiency of the motor driving the pump, and η_P is the pump efficiency. With total dynamic head of 30 m, flow rate of 6 m³/h, motor efficiency of 0.9 and pump efficiency of 0.7, the required power input, including a 20 % safety margin, is 1 kW. This corresponds 0.18 kWh/m³ of raw water.

- The power requirement for raw water treatment, including fine screening, coagulation/filtration + reverse osmosis is 3 kWh/m³ of raw water, based on Simoes et al. (2021).
- The power requirement for waste disposal (sludge pumping after treatment) is 0.26 kWh/m³ of raw water (Simoes et al., 2021).

Based on the above, the total energy consumption of water collection and transport, water treatment, and disposal of water treatment waste is 3.44 kWh/m^3 of raw water (=0.06 kWh/kg of hydrogen produced).

Table 4 presents the investment costs applied to raw water collection, treatment and storage systems, including installation, terrain preparation etc. The cost data mainly follows Simoes et al. (2021). The distance between the water source and the H_2 production plant was set as 1 km.

Table 4. Investment cost for water collection, treatment and storage

Process/equipment	Specific	Total €
Raw water collection (pipes, pump, accessories)		50,000
Raw water transport (stainless steel piping) 1 km	115 €/m	115,000
Raw water treatment equipment		
- fine screening (2,5 m x 5 m)		50,000
 filtration/coagulation (2 m x 5 m) 		40,000
- reverse osmosis (compact membranes, 6 m x		250,000
Disposal of water treatment waste		150,000
Water storage tank, 700 m ³	437 €/m³ *	305,000
Total		960,000

* 230 €/m³ times a factor of 1.9 to account for installation costs

3.5 Water management in the electrolysis process

The water management inside the electrolyzer includes the following subsystems:

- array level anodic circulation loops consisting of recirculation pumps, ion-exchangers, and gas/water separator vessels, and
- array level cathodic circulation loops consisting of gas/water separator vessels.

3.5.1 Anodic circulation loop

Electrolysis stacks are supplied with water through an array-specific anode water circulation system consisting of a water circulation pump, an ion exchanger, and an oxygen/water separator vessel.

Circulation pump

In a PEM electrolyser system, the circulating water pump is one of the key equipment. The circulating water provides a steady supply of reaction water to the anode for the electrochemical reaction, removes oxygen bubbles generated at the anode, and, most importantly, extracts the heat produced by internal dissipations and controls the cell temperature (Villagra & Millet, 2019). Indeed, large amounts of heat are released during electrolysis, and in order to maintain the cell temperature within an acceptable limit and avoid the formation of thermal gradients that can cause membrane swelling, mechanical stress and performance degradation in the catalyst layers (Villagra & Millet, 2019), very large water flow rates are required. Typically, the temperature increase must be kept below 5 °C.

There are two methods for internal cooling. The electrolyser can be cooled either by an excess amount of process water, or by separate cooling circuits inside the bipolar plates (Tiktak, 2019). Unfortunately, for the latter case, the design parameters of the flow fields are kept confidential by the manufacturers, and no reference information is available in the literature. Therefore, this study assumed that the process cooling is solely performed by an excess amount of process water.

The volumetric water flow rate for cooling, i_{H2O} in m³/s, pumped through a single stack array (10 MW) was calculated as follows (Villagra & Millet, 2019):

$$i_{H2O} = \frac{j * S_{EC}^{tot} * (U^{cell} - U^{tn})}{C_{H2O}^{p} * \rho_{H2O} * \Delta T}$$

where

- *-j* is the current density 2 A/cm²,
- S_{EC}^{tot} is the total electrochemical surface area of a single stack array 2,640,000 cm²

- U^{cell} is the nominal cell voltage at end-of-life 2.06 V
- U^{tn} is the thermoneutral voltage 1.48 V
- C_{H20}^{p} is the heat capacity of liquid water at constant pressure 4.187 x 10³ J/kgK
- ρ_{H20} is the water density 1000 kg/m³
- ΔT is the water temperature difference between cell outlet and inlet in Kelvin

The maximum temperature increase was set to 5 °C. When reaction water was added (10 kg/kg H_2), the total volumetric flow rate of water pumped through a single stack array at end-of-life reaches 529 m³/h, which was used as the basis for dimensioning the circulation pump.

With a total dynamic head of 80 m, a maximum flow rate of 529 m³/h, a motor efficiency of 0.9, and a pump efficiency of 0.7, the required power input of a single pump, including a 10 % safety margin, is 200 kW.

The investment cost for a single array-specific circulation pump is $31,000 \in$, based on the supplier's quotation. So, the total cost of two circulation pumps is $62,000 \in$.

Ion-exchanger

For electrolysis systems operating on ultrapure water, water treatment does not stop with the makeup water. It is also necessary to continuously purify (or polish) the ultrapure water that recirculates within the electrolyzer. After entering the electrolyzer system, the water is continuously contaminated with metal ions from the piping and process equipment (Madsen, 2022). Without countermeasures, the water quality in the circulation loop would quickly decrease and lead to system degradation.

To solve this problem, a side stream polisher is installed in the anode water circulation loop. The two processes available for handling the side stream polishing are mixed bed ion exchange and electro-deionization (EDI) (Madsen, 2022). The advantage of electro-deionization is that it is a chemical-free process, and the only consumable is DC power. Unlike mixed bed technology, EDI is a continuous process that does not require interrupting the operation to regenerate ion exchangers. Thus, EDI also eliminates the need to store and handle hazardous chemicals used for resin regeneration in mixed beds and does not produce a hazardous waste stream.

The side stream loop should be sized to remove contaminants at a rate equal to or higher than the release rate of contaminants to avoid accumulation in the electrolyzer. Typically, the polishing loop flow is 2-10 % of the anode circulation flow rate (Madsen, 2022). The higher the percentage, the higher the quality of the water sent to the electrolyzer.

The polishing loop flow was at 10 % of the anode circulation flow. The power consumption of EDI was set at 0.28 kWh/m³ of process water, based on Tseng et al. (2022).

The investment cost data for ion exchangers were retrieved from Fedorenko (2004). The investment cost of one EDI unit is 180,000 \in_{2023} (in total 360,000 \in).

Oxygen/water separator

Oxygen and water coming out of the stack are separated in array-level gas/water separators. From the gas/water separator, the water continues its flow to the recirculation phase, and the oxygen gas leaves the system through a demister, which removes fine water droplets from the oxygen stream.

The investment cost of the oxygen separator was set at $360,000 \in_{2023}$ per stack array (720,000 \in in total), based on Ruth et al. (2017). In addition, $160,000 \in$ is added to the investment costs of the anode-side water circulation loops to cover the cost of piping, fittings, valves, and measurement and instrumentation devices (flow meters, pressure sensors and temperature sensors).

3.5.2 Cathodic circulation loop

In most PEM electrolysis systems, the cathode side does not have its own water circulation. However, due to a drag current, water enters from the anode side to the cathode side. For this reason, a smaller gas-water separator is also installed on the cathode side. The water that has accumulated in the gas-water separator is returned to the anode side via a drain valve. (Holst et al., 2021)

Investment cost data for hydrogen/water separators were also retrieved from Ruth et al. (2017) and indexed to \in_{2023} currency. Hydrogen/water separator cost for a singe 10 MW stack array is 150,000 \in , so the total cost for two array-specific separators is 300,000 \in . In addition, a total of 100,000 \in is added to the investment costs of the cathode-side circulation loops to cover the cost of hoses, valves, instrumentation and controls.

3.6 Gas purification subsystem

Hydrogen gas produced by the electrolyser stack carries traces of impurities and is further treated to achieve the desired hydrogen purity. In the first step, small traces of oxygen, which crossed over the membrane into the hydrogen product gas stream, are removed in a deoxidation reactor, usually based on a palladium catalyst. On this catalytic surface, residual oxygen (ppm level) reacts with hydrogen to form water. The reaction is strongly exothermic, which leads to a considerable temperature increase of the hydrogen gas. To cool the gas stream and remove most of the water content from the hydrogen stream, a cooling system is installed after the deoxidiser. To lower the dew point of hydrogen to a level required for most applications, refrigeration system designed to achieve ultra-low temperatures is expected. For final drying, pressure swing adsorption is used. A schematic layout of the hydrogen purification unit is shown in Figure 9.



Figure 9. Hydrogen purification unit layout (Holst et al., 2021)

3.6.1 Deoxidation system

The presence of oxygen impurity in hydrogen can cause a potential safety risk when H₂ is used as a power source. Therefore, oxygen must be removed from the H₂ flow to meet the strict purity requirements (Xue et al., 2024). The deoxidation reaction is carried out in a plug flow reactor. The catalyst used is Pd (0.5wt-%) on alumina support. On this catalytic surface, the oxygen reacts with hydrogen to produce water vapour ($O_2+2H_2->2H_2O$), which is subsequently removed by drying processes. Deoxidation leads to the first hydrogen loss within the purification unit.

The deoxidation system sizing and costing are based on the HyJack Sizing and Costing online calculation tool (HyJack, 2022). The main inputs of the calculator were hydrogen flow 393 kg/h, H₂ purity 99.75 %, and O₂ content 0.25 %. The target O₂ content was set at 0 ppm. The reactor size indicated by the sizing tool is 2000 x 93.5 mm. The output hydrogen mass flow is 391.9 kg/h, H₂

purity 99.719 %, and H_2O content 0.281 %. The outlet temperature of hydrogen after the exothermic deoxidation reaction is 100 °C.

The indicative total investment cost for the deoxidation system is 68,000 € (HyJack, 2022).

3.6.2 Cooling and condensation

The deoxidiser is followed by a cooler that removes most of the water content by decreasing the temperature and condensing the water vapour. The aqueous water is then separated and removed from the process by a condensate trap.

The required dew point (= the temperature where water vapor condenses into aqueous water) of hydrogen in industrial applications depends on the specific requirements of the application. Lower dew points generally indicate drier hydrogen gas, which is desirable for applications where moisture can have detrimental effects. The typical dew point requirement in, e.g., fuel cell applications, hydrogenation reactions used in chemical production, and ammonia synthesis is between -60 °C and -20 °C.

Lowering the dew point of hydrogen to such low temperatures requires a specialized refrigeration system. These systems typically cool gases using a combination of compression, expansion, and heat exchange processes. This study applies a cascade system consisting of two basic compression refrigeration cycles – a high-temperature cycle (HTC) and a low-temperature cycle (LTC) – operating at progressively lower temperatures to achieve the desired dew point for the hydrogen gas. The LTC cycle and the HTC cycle are connected via a cascade heat exchanger (Figure 10). The high-temperature cycle uses a refrigerant capable of cooling the gas stream down to a moderate temperatures. In this study, the evaporation temperature was assumed to be -50 °C. Refrigerants are selected based on their boiling points, which directly affect the temperature range over which they can effectively absorb or release heat. The primary advantage of cascades systems is that the two cycles do not necessarily contain the same refrigerants: In the low-temperature side, a refrigerant with a higher vapour pressure can be used, while a refrigerant with a lower vapour pressure is suitable for the high-temperature cycle. (SWEP, 2022; LAC, 2024)



Figure 10. Schematic illustration of cascade cooling system (Koşan, 2023)

Cascade refrigeration systems are known for their high energy efficiency compared to single-stage refrigeration systems, especially for achieving very low temperatures. By optimizing the temperature levels of each cooling stage and minimizing heat losses, cascade cooling systems can effectively dry hydrogen gas while minimising energy consumption.

The amount of heat (Q) that needs to be removed through the cascade system is 140 kW. The power consumption of the cascade refrigeration system was determined based on its coefficient of performance (COP) value:

$$W = \frac{Q}{COP}$$

Applying a COP value of 1.23 (Zheng et al., 2023), the power consumption of the system is 0.29 kWh/kg of hydrogen.

The capital cost per kW of refrigeration load for this 2-stage system was assumed to be 4,200 €₂₀₂₃, based on Luyben (2017). Thus, the total cost of the cascade refrigeration system considered in this study is 588,000 €.

After the cascade refrigeration, part of the water remains present. Thus, further drying step is required. The final drying is based on pressure swing adsorption (PSA), a cyclic process that uses solid adsorbent beds to remove impurities from the gas. During the adsorption step, the hydrogen gas stream containing traces of water is passed through the adsorption material layer at elevated pressure. The water molecules are selectively adsorbed on the surface of the adsorbing particles, leaving pure hydrogen gas to exit the system. After the adsorption step, the adsorption layer becomes saturated with moisture and must be regenerated to remove the trapped water molecules. This is achieved by reducing system pressure. By cycling between the adsorption and regeneration stages, PSA effectively removes moisture from the hydrogen gas stream, resulting in a purified product gas with practically zero moisture content. For continuous drying, two columns are required; while the first vessel is drying in-line the hydrogen stream, the second one is undergoing regeneration off-line (Holst et al., 2021).

A typical PSA hydrogen drying system consists of adsorbent vessels packed with desiccant, valves, control systems, and instrumentation to regulate pressure fluctuations and monitor system performance. The investment cost of PSA applied in this study was retrieved from the HyJack Sizing and Costing online calculation tool (HyJack, 2022). The total cost for equipment and auxiliaries, including engineering, transportation, instrumentation and piping, is 946,000 €.

The recommended operating conditions are 40 bar and 30 °C for the adsorption and 0.1 mbar and 30 °C for regeneration. The power consumption of the system is 0.50 kWh/kg of hydrogen.

The adsorbent chosen is Molecular Sieves 4A, and the mass of adsorbents per column is 5,774 kg. The PSA modelled decreases the water content to <1 ppm. The amount of H_2 leaving the PSA is 390.8 kg/h with a purity of 99.9998 %.

A further treatment of oxygen gas produced on the anode side is not considered to be within the scope of this paper. Thus, oxygen is simply vented into the atmosphere. However, it should be kept on mind that the effective utilization of oxygen byproduct would be one strategy to cut down the hydrogen production costs.

3.7 Hydrogen compression and storage

The density of hydrogen under standard conditions is very low, 0.08988 kg/m³, resulting in a low volumetric energy density. For storage, transportation, and further processing, a final pressure of 80 to 900 bars is required (Holst et al., 2021)

The compression of hydrogen is a state-of-the-art process used in nearly all hydrogen-related applications. For hydrogen compression in large-scale applications, reciprocal, multi-stage compressors are typically used (Holst et al., 2021).

Hydrogen compression and storage were considered to be beyond the scope of this paper.

3.8 Thermal management subsystem

In electrolysis, about one-third of the applied electrical energy is released in the form of heat. This heat must be removed from the system to maintain the correct operating temperature and to avoid cell damage. The relationship among electric power input $P_{el_{f}}$ heat produced Q_{el} and hydrogen energy P_{H2} is as follows:

$$\begin{split} P_{el} &= U^{cell} I^{cell} N^{cell} \\ Q_{el} &= (U^{cell} - U^{th}) I^{cell} N^{cell} \\ P_{H2} &= P_{el} - Q_{el} \end{split}$$

The design capacity of the cooling system in this study was dimensioned based on the end-of-life situation. As already stated, a lifetime of 80,000 hours results in voltage degradation, which leads to an increase in the amount of heat production if the hydrogen production rate is kept constant. The total amount of heat that needs to be dissipated from the system at end-of-life is 6,125 kW.

For process cooling, a heat exchanger is placed in both array-specific anodic water circulation loops. In addition to process water, product gases are cooled to condense water from the gas flows. The condensed water is then fed back to the electrolysis system to reduce water demand.

In this study, the cooling water circuit is designed to be implemented as closed loop. The closedloop cooling water circuit is used to cool all the different heat exchangers in the hydrogen production process. Heat is extracted from the stacks from three different sources: the water recirculation loops, the oxygen streams, and the hydrogen streams (Figure 11). Cooling of the central cooling water circuit is realised with dry coolers.



Figure 11. The cooling system of an electrolyzer stack. The cooling water flow is shown in red and blue.

In case the excess heat would be used by a third party, the design has to be adapted to make this possible with an extra tie-in on the cooling system.

3.8.1 Heat exchangers

For the anode circulation loops, compact thermally efficient gasketed plate heat exchangers were selected. The heat exchanger data were defined using the Kelvion online Select PHE tool (Klevion, 2024). Table 5 presents the main parameters of the anode circuit heat exchangers.

Table 5. Anode circuit heat exchangers

	Hot side	Cold side	
Heat exchanged	30	kW	
Fluid	water	water	
Volume flow	529	175	m³/h
Inlet temp.	70	40	°C
Outlet temp.	65	55	°C
Pressure drop	246	32	mbar
Connections	DN		

The capital cost of anode circuit heat exchangers is $39,000 \in$ each (in total $78,000 \in$).

In addition, hydrogen and oxygen streams are cooled to condense water from the gas flows. For cooling of the gas streams, gas-to-water brazed plate heat exchangers are used. The product gases are cooled down to 41 °C. The amount of heat Q (in kW) that needs to be dissipated from the gas flows was determined as follows:

$$Q = \dot{m} * C_p * \Delta T$$

where \dot{m} is the mass flow of gas, C_{ρ} is the specific heat of gases, and ΔT is the temperature difference between gas inlet and outlet in °C.

The amount of heat that needs to be dissipated from the hydrogen flow per one stack-array is 23 kWh (hydrogen mass flow 196.5 kg/h, c_p 14.3 kJ/kg °C, ΔT 29 °C). The amount of heat to be transferred from the oxygen flow per one stack array is 12 kWh (oxygen mass flow 1,572 kg/h, c_p 0.918 kJ/kg °C, ΔT 29 °C). So, a total of 70 kWh heat is dissipated from gas flows.

Investment costs for gas-to-water heat exchangers were calculated based on the specific cost of $500 \in \text{per m}^2$ of heat transfer surface area (A). A was calculated from the equation:

$$A = \frac{Q}{U * LMTD}$$

where U is heat transfer coefficient in $W/m^{2\circ}C$, and the LMTD is a logarithmic average of the temperature difference between the hot and cold feeds at each end of heat exchanger and is calculated using the formula:

$$LTDM = \frac{(T_1 - T_4) - (T_2 - T_3)}{\ln\left[\frac{(T_1 - T_4)}{(T_2 - T_3)}\right]}$$

where T_1 is the temperature at hot inlet, T_2 the temperature at hot outlet, T_3 the temperature at cold inlet, and T_4 the temperature at cold outlet.

Assuming a U-value of 140 W/m²°C (Dimian & Bildea, 2008) for gases, the total capital cost of gasto-water heat exchangers is 48,000 €.

3.8.2 Dry cooler

A dry cooler is a type of heat exchanger used to cool fluids without the need for water as a cooling medium. Unlike traditional cooling towers or evaporative coolers, which use water evaporation to remove heat, a dry cooler uses ambient air to cool the working fluid.

The dry cooler consists of a series of finned tubes made of materials with high thermal conductivity, such as copper or aluminium. These tubes are designed to maximise the surface area available for

heat transfer. The fans, mounted near the heat transfer surface, pull the ambient air and circulate it over tubes containing cooling water. As the ambient air passes through the system, heat is transferred from the fluid inside the tubes to the air through convection. The heated air leaves the dry cooler, while the cooled water is circulated through heat exchangers attached to associated processes.

The total cooling load of the dry coolers is 6,125 kW. To define the dry cooler specifications, the Kelvion online Select RT tool was used (Kelvion, 2024). The central cooling system consists of 4 dry cooler units, 1530 kW each. The main specifications of dry coolers are presented in Table 6.

Table 6. Main specifications of dry coolers

Parameter	Value (per unit)
Capacity	1,530 kW
Volume flow rate	87.7 m³/h
Fluid inlet	55 °C
Fluid outlet	40 °C
Operation mode	AC
Fans	12 pieces
Power input per fan	1,650 W
Pressure drop	276 kPa

The energy consumed by dry coolers is only fan power. The power requirement varies depending on outdoor temperatures; the lower the ambient temperature, the lower the power consumption. The frequencies of the temperature ranges applied in this study and the corresponding power consumptions are presented in detail in Appendix 1. The annual average power input per unit is 8.0 kW (32 kW in total).

The capital cost of one dry cooler unit, including wiring and connections, is 72,000 €, and the total cost of 4 units is $288,000 \in$.

3.8.3 Cooling water pump

At first, the required cooling water flow to dissipate 6,125 kW of heat from the system was calculated based on the formula:

$$\dot{\mathbf{m}} = \frac{Q}{c_p \left(T_{out} - T_{in}\right)}$$

With a total dynamic head of 90 m, flow rate of 351 m³/h, pump efficiency of 0.7, and motor efficiency of 0.9, the required power input of the cooling pump is 135 kW. The capital cost of the cooling pump was estimated at $60,000 \in$.

In addition, the cooling system includes safety features to prevent issues such as water leakage, overcooling, or system failure. Pressure sensors, flow meters, and temperature alarms are commonly used to monitor the cooling system and ensure safe operation. The investment cost for instrumentation was set at $50,000 \in$. Moreover, $20,000 \in$ was added for piping (Ruth et al., 2017).

3.9 Safety systems

Hydrogen gas is highly flammable, so safety is paramount in hydrogen production units. Combustible gas detectors are used to trigger emergency shutdowns in case of leaks to avoid hazardous conditions in the plant. The capital cost of gas detectors and exhaust ventilation was set at 100,000 euros.

4 MATLAB/Simulink modelling

To reduce the complexity, only the electrolyzer model is considered (raw water treatment and gas purification excluded). The real output of the system is only the useful hydrogen produced. Oxygen is not viewed as a real output of the system; however, it is also produced along with hydrogen.

The proposed PEM electrolysis model on MATLAB/Simulink, shown in Figure 12, includes the following main subsystems:

- Electrical supply system
- Water supply
- Electrolyzer
- Anodic circulation loop
- Cathodic circulation loop
- Cooling system
- Hydrogen output

The block represents the electric load for a PEM electrolysis system, calculating hydrogen generation based on water temperature and electrical energy provided.



Figure 12. System-level block diagram model of PEM electrolysis implemented in MATLAB/Simulink

4.1 Simulation results for the proposed PEM electrolysis block system

The simulation was performed across a current density range of 0.01–2.0A/cm². The following base parameters were used:

- Reversible voltage: 1.23 V
- Ohmic resistance: 0.2 Ω·cm²
- Activation overpotential coefficient: 0.03 V
- Concentration overpotential coefficient: 0.05 V
- Operating temperature: 60–80°C
- Operating pressure: 1–10 bar

Hydrogen output was calculated from Faraday's law, assuming 100% efficiency in the ideal case. Power output was computed as the product of voltage and current density.

Current-voltage (I–V) and current-power (I–P) curves were generated to assess performance (Fig. 13). The I–V curve demonstrated a typical rising profile with increasing current density, while the I–P curve peaked before efficiency losses set in at higher currents. These trends confirm correct electrolyzer behavior.



Figure 13. I–V Curve (left): Voltage increases with current, as expected in an electrolyzer due to activation, ohmic, and concentration overpotentials. I–P Curve (right): Power increases with current density, showing no peak within the simulated range, consistent

Figure 14 illustrates the power produced, consumed (kW), and heat power dissipated (kW) by the PEM electrolysis system.



Figure 14. Power produced, consumed (kW), and heat power dissipated (kW) by the PEM electrolysis system.

The plotted data show the distribution of input electrical power into two major outputs—chemical power (stored in hydrogen) and thermal losses (heat dissipation)—across the working range of the PEM electrolyzer block model. At lower power levels, most of the incoming electrical energy is successfully transformed into chemical energy, demonstrating good system efficiency. This is demonstrated by the near alignment of the power consumed and chemical power produced curves in the lower range. This region exhibits little heat power dissipation, indicating good thermal management and negligible activation/ohmic losses. As input power grows, particularly beyond midrange operation, the difference between power spent and chemical power output widens. This divergence correlates to an increase in heat power dissipation, implying an increase in system inefficiency, principally due to higher overpotentials and thermal resistance inside the membrane-electrode assembly.

The trend demonstrates that, while the PEM electrolyzer block diagram is highly efficient at low-tomoderate loads, performance suffers slightly at larger currents due to internal heat generation and increased resistance. This stresses the significance of efficient thermal management and control systems for achieving high efficiency under fluctuating load situations.

Figure 15 illustrates the stack temperature (°C) and hydrogen output (Nm³/h) graphs for the PEM electrolyzer block diagram.



Figure 15. Stack Temperature vs. Time (left): The temperature increases from the initial 60 °C and asymptotically stabilizes around 70 °C due to thermal control, simulating realistic heat accumulation and dissipation dynamics. Hydrogen Output Rate vs. Time (right

The system efficiency vs. input power plot for the proposed PEM electrolyzer model is shown in Fig. 16. The efficiency peaks (\sim 60%) at moderate power levels and declines to \sim 45% at full input (95 kW), matching the thermal and electrochemical losses described in the study.



Figure 16. System efficiency (%) vs. input power (kW)

Figure 17 presents the dynamic response and control validation results of the PEM electrolyzer system.



Figure 17. Dynamic response and control validation of PEM electrolyzer system

The graphic confirms the dynamic energy balance of the PEM electrolyzer system. The alignment between power utilized and chemical power produced at mid-range current densities demonstrates that electrical energy is converted into hydrogen efficiently and effectively. The low heat power dissipation recorded under steady-state settings implies that the thermal management subsystem is effective. These patterns reflect the simulation model's accuracy and realism, indicating its applicability for system-level design analysis and operational optimization of PEM electrolyzers. The modeled behavior is consistent with theoretical expectations and reported experimental standards, demonstrating model validity.

4.2 Discussion

The PEM electrolyzer system model was simulated under different input power settings to evaluate its dynamic behavior, hydrogen generation rate, efficiency, and thermal properties. The model contains detailed representations of the electrochemical cell stack, water feed system, gas separation and venting, thermal management, and DC power supply integration.

4.2.1 Stack performance and current density response

The model exhibits a constant electrochemical response across a power input range of 20 kW to 95 kW. At peak input power (95 kW), the electrolyzer operates at a current density of 1.25 A/cm², resulting in a cell voltage of about 1.85 V. The I-V and polarization curves created are consistent with theoretical and reported experimental data, supporting the model's ability to capture voltage losses due to activation, ohmic, and concentration overpotentials.

4.2.2 Hydrogen production rate

Hydrogen generation was computed using Faraday's law and the stack current. At full load, the system generates around 0.045 mol/s of hydrogen, equivalent to about 1.01 Nm³/hour. The model accurately follows the dynamic change in hydrogen output as input power varies, indicating that it is suitable for grid-connected or renewable-integrated operation.

4.2.3 Thermal behavior and cooling requirements

Heat generated from electrochemical reactions and resistive losses causes stack temperatures to rise fast under full load. Inadequate thermal management can cause stack temperatures to exceed 80 °C, resulting in reduced efficiency. The model optimizes membrane hydration and ionic conductivity by maintaining an operating temperature of 60-70 °C using active thermal control and a water-based cooling loop. Under full load conditions, the cooling requirement reaches a peak of approximately 28 kW.

4.2.4 System efficiency

The system's thermodynamic (higher heating value, HHV) efficiency peaks at around 60% under moderate load before dropping to 45% at full power due to increased ohmic losses and thermal load. The model discovers optimal operating points for balancing hydrogen production rate and energy efficiency, driving real-time power allocation algorithms.

4.2.5 Dynamic response and control validation

Dynamic tests, which included power step inputs and load transients, proved the integrated control system's stability. PID controllers for flow rate, pressure, and temperature showed good tracking performance with minimum overshoot. The model is resilient to quick changes, which may occur with renewable energy sources such as wind or solar input.

4.3 Conclusion

The developed MATLAB/Simulink model of a PEM electrolyzer block system has great fidelity and is useful for assessing system-level behavior under dynamic operating conditions. The model accurately reflects critical electrochemical characteristics, such as voltage losses and current density response, over a broad input power range (20-95 kW). Simulations show a high link between input power and hydrogen generation, with a maximum output of around 1.01 Nm³/h at full load. Proper thermal management is crucial for maintaining ideal operating temperatures (60-70°C) and avoiding

efficiency losses caused by overheating. The system reaches a peak thermodynamic efficiency of 60% under moderate loads, which drops to 45% at higher power levels. Finally, dynamic response tests show the control strategy's robustness, with PID controllers preserving system stability during power transients, indicating the model's appropriateness for integration with fluctuating renewable energy sources. These findings help with design optimization and real-time control development for sophisticated hydrogen production systems.

The MATLAB scripts employed in the modelling are documented in Appendix 2.

5 Economic analysis

The costs associated with green hydrogen production via electrolysis are influenced by three key factors: the capital cost of the electrolyser systems, the cost of renewable electricity, and the capacity factor of the hydrogen production plant.

In this study, two approaches for the connection of the electrolyser directly to the wind power generation were considered:

- In Scenario 1, the 20 MW electrolyser is connected to a 100 MW wind farm and only uses renewable electricity. Excess electricity is fed to the grid.
- In Scenario 2, the 100 MW wind farm primarily provides electricity to the grid and drives the electrolyser only during excess production. Therefore, the electrolyser operates on a more intermittent basis. The grid export threshold was set at 30 MW.

For comparative purposes, in Scenario 3, electrolysis plant is connected to the grid and uses grid power. In this case, hydrogen is not obtained with 100 % clean energy, and is not considered green.

The economic analysis is based on the LCOH (Levelized Cost of Hydrogen) model. The levelized cost of hydrogen represents the average cost of producing one kilogram of hydrogen over the lifetime of the production facility. It accounts for both capital expenditures and operational expenditures, including investment, maintenance, energy, and other associated costs.

Figure 18 provides a breakdown of the key cost components that contribute to the calculation of the LCOH.



Figure 18. Cost components included in the economic analysis

The economic analysis in this study builds upon the theoretical 20 MW PEM electrolyzer model detailed in Chapter 3.

5.1 Capital costs

The capital costs (CAPEX) of the electrolyser system comes from the sum of the stack and balance of plant (BOP) cost, installation, and indirect capital cost, such as engineering and design, site preparation, permitting and contingency. The installation costs, including transporting, placing, connecting and commissioning, were added to the total purchased equipment cost with a 14 % markup, based on Acevedo et al. (2023). Engineering cost was set at 10 % of the total installed capital cost (Badgett et al., 2024). Costs for site preparation and general facilities (site foundation, roads, buildings) were assumed at 2 % of the total installed capital cost, in line with the LCRI Hydrogen Electrolysis Techno-Economic Analysis (TEA) Tool (LCRI, 2024). Permitting cost, including legal and contractor fees cost, was set at 15 % of the total installed capital cost, and another 15 % was added to cover the uncertainty in the project cost estimate itself (cost contingency), based on LCRI (2024) and Badgett et al. (2024). The capital costs are summarized in Table 7.

12 340 000	€	
4 540 000	€	
960 000	€	
1 302 000	€	
400 000	€	
1 602 000	€	
544 000	€	
100 000	€	
21 788 000	€	1089 €/kW
3 050 320	€	14 % of total purchased equipment cost
24 838 320	€	
2 483 832	€	10 % of total installed capital cost
496 766	€	2 % of total installed capital cost
3 725 748	€	15 % of total installed capital cost
3 725 748	€	15 % of total installed capital cost
35 270 414	€	1764 €/kW
	12 340 000 4 540 000 960 000 1 302 000 400 000 1 602 000 544 000 3 050 320 21 788 000 3 050 320 24 838 320 24 838 320 24 838 320 3 725 748 3 725 748	12 340 000 € 4 540 000 € 960 000 € 1 302 000 € 4 400 000 € 400 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 1 602 000 € 3 050 320 € 2 483 8320 € 3 725 748 € 3 725 748 € 3 725 741 €

Table 7. Total installed capital cost

5.2 Operating costs

Operating expenses (OPEX) consist of two main elements: a fixed OPEX component and a variable OPEX component. Fixed operating costs consist of operating and maintenance costs, including operating labour, routine maintenance (excluding replacements) and overhead. In this study, fixed O&M costs were estimated at 2 % of direct capital costs, in line with LCRI (2024) and KPMG (2022). Variable OPEX mainly consists of electricity purchase costs. In addition to the electrolysis process, minor electricity consumption is associated with, e.g., water treatment, process cooling and gas purification processes. The average total power consumption of the electrolyser (stack + BOP) in Scenario 1 and 3 is 54.8 kWh/kg of hydrogen, and in Scenario 2, 54.6 kWh/kg H₂. With annual operating hours of ~8000 h (scenarios 1 and 3), the stack is replaced after 10 years of service life. In scenario 2, with lower annual operating hours, the stack replacement takes place later, after 15 years of service life. More detailed information of specific power consumption of the electrolysis system in different scenarios is given in Table 8.

Table 8. Specific power consumption of the electrolysis system. Upper table for Scenarios 1 and 3, and the lower table for Scenario 2.

Year	Stack	power		BOP power consumption							
	consu kWh/k	mption (a of H2		kWh/kg of H2							
				Water collection	Water Anodic circulation loop Hydrogen purification The				Thermal m	anagement	
	Ucell	Stack power	Amount of circulating water m ³ /h	Water collection, treatment and storage	Circulating pump	EDI	H ₂ cooling and condensation	PSA	Cooling water pump	Dry coolers (fan power)	BOP total
1	1.908	51.6	781	0.06	0.88	0.06	0.29	0.50	0.35	0.08	2.21
2	1.916	51.8	796	0.06	0.89	0.06	0.29	0.50	0.35	0.08	2.23
3	1.924	52.0	811	0.06	0.91	0.06	0.29	0.50	0.35	0.08	2.24
4	1.932	52.2	825	0.06	0.92	0.06	0.29	0.50	0.35	0.08	2.26
5	1.940	52.4	840	0.06	0.94	0.06	0.29	0.50	0.35	0.08	2.28
6	1.948	52.7	855	0.06	0.96	0.06	0.29	0.50	0.35	0.08	2.30
7	1.957	52.9	869	0.06	0.97	0.06	0.29	0.50	0.35	0.08	2.31
8	1.965	53.1	884	0.06	0.99	0.06	0.29	0.50	0.35	0.08	2.33
9	1.973	53.3	899	0.06	1.01	0.06	0.29	0.50	0.35	0.08	2.35
10	1.981	53.5	913	0.06	1.02	0.07	0.29	0.50	0.35	0.08	2.37
				St	ack replace	nent					
11	1.908	51.6	781	0.06	0.88	0.06	0.29	0.50	0.35	0.08	2.21
12	1.916	51.8	796	0.06	0.89	0.06	0.29	0.50	0.35	0.08	2.23
13	1.924	52.0	811	0.06	0.91	0.06	0.29	0.50	0.35	0.08	2.24
14	1.932	52.2	825	0.06	0.92	0.06	0.29	0.50	0.35	0.08	2.26
15	1.940	52.4	840	0.06	0.94	0.06	0.29	0.50	0.35	0.08	2.28
16	1.948	52.7	855	0.06	0.96	0.06	0.29	0.50	0.35	0.08	2.30
17	1.957	52.9	869	0.06	0.97	0.06	0.29	0.50	0.35	0.08	2.31
18	1.965	53.1	884	0.06	0.99	0.06	0.29	0.50	0.35	0.08	2.33
19	1.973	53.3	899	0.06	1.01	0.06	0.29	0.50	0.35	0.08	2.35
20	1.981	53.5	913	0.06	1.02	0.07	0.29	0.50	0.35	0.08	2.37
Average		52.5		0.06	0.95	0.06	0.29	0.50	0.35	0.08	2.29

Year	Stack	power		BOP power consumption							
	consu kWh/k	mption a of H ₂		kWh/kg of H2							
	KUII/I			Water collection	Anodic circula	tion loop	Hydrogen purif	fication	Thermal m	anagement	
	Ucell	Stack power	Amount of circulating water m ³ /h	Water collection, treatment and storage	Circulating pump	EDI	H ₂ cooling and condensation	PSA	Cooling water pump	Dry coolers (fan power)	BOP total
1	1.905	51.5	0.06	776	0.87	0.06	0.29	0.50	0.35	0.08	2.20
2	1.911	51.6	0.06	786	0.88	0.06	0.29	0.50	0.35	0.08	2.21
3	1.916	51.8	0.06	796	0.89	0.06	0.29	0.50	0.35	0.08	2.22
4	1.922	51.9	0.06	806	0.90	0.06	0.29	0.50	0.35	0.08	2.24
5	1.927	52.1	0.06	816	0.91	0.06	0.29	0.50	0.35	0.08	2.25
6	1.933	52.2	0.06	826	0.92	0.06	0.29	0.50	0.35	0.08	2.26
7	1.938	52.4	0.06	836	0.93	0.06	0.29	0.50	0.35	0.08	2.27
8	1.944	52.5	0.06	846	0.95	0.06	0.29	0.50	0.35	0.08	2.28
9	1.949	52.7	0.06	856	0.96	0.06	0.29	0.50	0.35	0.08	2.30
10	1.955	52.8	0.06	866	0.97	0.06	0.29	0.50	0.35	0.08	2.31
11	1.960	53.0	0.06	876	0.98	0.06	0.29	0.50	0.35	0.08	2.32
12	1.965	53.1	0.06	885	0.99	0.06	0.29	0.50	0.35	0.08	2.33
13	1.971	53.3	0.06	895	1.00	0.06	0.29	0.50	0.35	0.08	2.34
14	1.976	53.4	0.06	905	1.01	0.06	0.29	0.50	0.35	0.08	2.35
15	1.982	53.6	0.06	915	1.02	0.07	0.29	0.50	0.35	0.08	2.37
			-	St	ack replacer	nent					-
16	1.905	51.5	0.06	776	0.87	0.06	0.29	0.50	0.35	0.08	2.20
17	1.911	51.6	0.06	786	0.88	0.06	0.29	0.50	0.35	0.08	2.21
18	1.916	51.8	0.06	796	0.89	0.06	0.29	0.50	0.35	0.08	2.22
19	1.922	51.9	0.06	806	0.90	0.06	0.29	0.50	0.35	0.08	2.24
20	1.927	52.1	0.06	816	0.91	0.06	0.29	0.50	0.35	0.08	2.25
Average		52.3	0.06		0.93	0.06	0.29	0.50	0.35	0.08	2.27

Specific power costs were determined as follows:

Scenario 1: In Scenario 1, the electrolyzer is powered by wind energy. The cost of this electricity is determined by the market value of wind power—that is, the prices at which electricity generated by wind farms is sold on the market at various times. Wind power typically has a lower market value than the average electricity price. This is primarily due to its intermittent and unpredictable nature, which limits its availability during periods of high electricity demand and peak pricing. Furthermore, when wind generation is high, the resulting increase in supply often drives market prices down, further reducing the revenue that wind power producers can capture.

The market value of wind power was calculated based on hourly wind power generation and hourly electricity market prices:

$$Market Value \left(\frac{\epsilon}{MWh}\right) = \frac{\sum (Wind \ power \ generation_{hour} * \ Market \ price_{hour})}{\sum Wind \ power \ generation_{hour}}$$

Hourly wind power production data were obtained from Finnish Energy (2024), and hourly electricity market prices are based on Nord Pool prices. The review period was 1 January 2023–31 December 2023. During this period, the total wind power production was 14,469,710 MWh, and the total revenue was 577,871,168 €, which results in the wind power market value of 40 €/MWh. As the electrolysis plant is directly connected to the wind farm, electricity transmission and distribution fees are avoided.

Scenario 2: In Scenario 2, the wind farm primarily supplies electricity to the grid. The electrolyzer operates only when more wind power is produced than can be fed into the grid. The price of this surplus electricity was determined based on the production cost of wind power, calculated using the levelized cost of electricity (LCOE) method:

$$LCOE = \frac{Investment \ cost + \sum_{i=1}^{n} \frac{Annual \ O\&M \ cost_{i}}{(1+r)^{i}}}{\sum_{i=1}^{n} \frac{Amount \ of \ electricity \ produced \ per \ year_{i}}{(1+r)^{i}}}$$

where n is the plant lifetime in years and r is the discount rate of the project, covering financial charges and return on investment.

The LCOE for a 100 MW wind farm was calculated using the following assumptions: Initial investment cost 1.3 M€/MW (Renewables Finland, 2024), O&M cost 2 % of the initial investment, capacity factor 42.3 %, inflation 2 %/year, discount rate 6 %, and lifetime 25 years. Under these assumptions, the resulting LCOE is 36€/MWh, which aligns well with reported figures for onshore wind power in Finland (IRENA 2023).

Scenario 3: In Scenario 3, the electrolysis plant is connected to the grid and only uses grid power. In this scenario, the electricity price is based on Finnish Energy's (2024) total electricity production data and Nord Pool's hourly electricity market prices. Again, the review period was 1 January 2023–31 December 2023. During this period, the total power production was 78,025,001 MWh and the total revenue was 4,386,691,548 €, which results in grid power price of 56 €/MWh. By adding a margin of 2 €/MWh and a distribution fee of 10 €/MWh (StatFin, 2024), the total cost of grid electricity amounts to 68 €/MWh.

The water source is surface water from the nearby lake, making the cost of water supply negligible.

Additional variable O&M cost consists of those required for stack replacement. For example, with 8,000 annual operating hours, the 80,000-hour durability (stack lifetime) is reached in 10 years, after which the cell stacks need to be replaced. In practice, the cost for stack replacement can either be included in the capital costs or presented as annual operating costs (total cost annualized over project life). To highlight the significance of stack replacement costs, we present it as a single cash flow that allows it to be directly compared with other cash flows. The stack replacement cost percentage is estimated at 31 % of the installed capital cost, based on LCRI (2024). This cost is intended to cover the net costs of replacing the stack including installation costs.

5.3 Other costs

Finance cost, also known as financial expense or interest expense, refers to the cost of financing the company's operations and investments through debt or other forms of financing.

Decommissioning costs means all costs and expenses related to decommissioning and removing the unit from service. For the sake of simplification, this study assumes that decommissioning costs are offset by the plant's salvage value (=residual value).

5.4 Levelized cost of hydrogen

The levelized cost of hydrogen (LCOH) is a metric used to evaluate the overall cost of producing hydrogen over the lifetime of a production facility. It is analogous to the levelized cost of energy (LCOE) used in the electricity sector. The LCOH accounts for all the costs associated with producing hydrogen and divides them by the total amount of hydrogen produced over the system's lifetime. The final result is an average cost of production per unit, expressed in \notin /kg. LCOH is calculated using the following formula:

$$LCOH = \frac{CAPEX + Stack replacement + \sum_{i=1}^{n} \frac{Power cost_i + 0\&M_i}{(1+r)^i}}{\sum_{i=1}^{n} \frac{Hydrogen output_i}{(1+r)^i}}$$

where n is the plant lifetime in years and r is the discount rate of the project, covering financial charges and return on investment.

The service life of the electrolysis plant was set at 20 years. Electricity prices were assumed to increase by 2 % per year. An annual price increase of 2 % was also applied to operating and maintenance costs. In the base case, a discount rate of 8 % was used, in line with the IEA (2020) and IRENA (2021) reports.

5.4.1 Scenario 1

In Scenario 1, electrolyzer is connected to a wind farm and only uses renewable electricity. Annual operating hours of the hydrogen production plant are 8,076 h (wind speed below the cut-off speed of 3.5 m/s for 334 h/a, electrolysis plant availability 96 %). Based on the historical wind data described in Chapter 2.5, the amount of wind power available for hydrogen production in Scenario 1 is 157,232 MWh/a. Annual hydrogen production slightly varies due to voltage deterioration over time, thus, hydrogen output varies between 3,050 and 2,938 t/a. Applying the wind power price of $40 \in /MWh$, the levelized cost of hydrogen in Scenario 1 is $4.1 \in /kg$. More detailed results are presented in Table 9. Electrolysis plant's capacity factor in Scenario 1 is 90 %.

Table 9. Levelized cost of hydrogen in Scenario 1. Wind power price 40 €/MWh, discount rate 8 %, annual price increase 2 %, stack lifetime 80,000 h.

Year	Wind power	Hydrogen	CAPEX	Power cost	Operation and	Stack
	hydrogen	output			cost	replacement
	production					
	MWh/a	kg	€	€	€	€
0	0	0	35 270 414	0	0	0
1	157 232	3 049 550	0	6 289 273	496 766	0
2	157 232	3 036 697	0	6 415 059	506 702	0
3	157 232	3 023 952	0	6 543 360	516 836	0
4	157 232	3 011 314	0	6 674 227	527 172	0
5	157 232	2 998 780	0	6 807 712	537 716	0
6	157 232	2 986 351	0	6 943 866	548 470	0
7	157 232	2 974 024	0	7 082 743	559 440	0
8	157 232	2 961 799	0	7 224 398	570 628	0
9	157 232	2 949 673	0	7 368 886	582 041	0
10	157 232	2 937 647	0	7 516 264	593 682	0
11	157 232	3 049 550	0	7 666 589	605 555	7 699 879
12	157 232	3 036 697	0	7 819 921	617 667	0
13	157 232	3 023 952	0	7 976 319	630 020	0
14	157 232	3 011 314	0	8 135 845	642 620	0
15	157 232	2 998 780	0	8 298 562	655 473	0
16	157 232	2 986 351	0	8 464 534	668 582	0
17	157 232	2 974 024	0	8 633 824	681 954	0
18	157 232	2 961 799	0	8 806 501	695 593	0
19	157 232	2 949 673	0	8 982 631	709 505	0
20	157 232	2 937 647	0	9 162 283	723 695	0
Total		59 859 574	35 270 414	152 812 796	12 070 117	7 699 879
NPV		29 462 324	35 270 414	71 403 444	5 639 894	7 699 879
LCOH	4.1	€/kg				

5.4.2 Scenario 2

In Scenario 2, the wind farm primarily supplies electricity to the grid and drives the electrolyzer only during excess production. Hence, the electrolyser operates on a more intermittent basis. Based on the historical wind data presented in Chapter 2.5, the annual operating hours of the hydrogen production plant in Scenario 2 are 5,335 h, and the amount of wind power available for hydrogen production is 92,692 MWh/a. Hydrogen output varies between 1,665 and 1,599 t/a. Applying the wind power price of 36 \in /MWh, the levelized cost of hydrogen in Scenario 2 is 5.3 \in /kg. More detailed results are presented in Table 10. Electrolysis plant's capacity factor in Scenario 2 is 51 %.

Table 10. Levelized cost of hydrogen in Scenario 2. Wind power price 36 €/MWh, discount rate 8 %, annual price increase 2 %, stack lifetime 80,000 h.

Year	Wind power	Hydrogen	CAPEX	Power cost	Operation	Stack
	available for	output			and maintenance	replacement
	production				cost	
	MWh/a	ka	€	€	€	€
0	0	0	35 270 414	0	0	0
1	89 420	1 665 489	0	3 219 102	496 766	0
2	89 420	1 660 564	0	3 283 484	506 702	0
3	89 420	1 655 668	0	3 349 154	516 836	0
4	89 420	1 650 800	0	3 416 137	527 172	0
5	89 420	1 645 962	0	3 484 460	537 716	0
6	89 420	1 641 151	0	3 554 149	548 470	0
7	89 420	1 636 369	0	3 625 232	559 440	0
8	89 420	1 631 614	0	3 697 736	570 628	0
9	89 420	1 626 887	0	3 771 691	582 041	0
10	89 420	1 622 188	0	3 847 125	593 682	0
11	89 420	1 617 515	0	3 924 068	605 555	0
12	89 420	1 612 869	0	4 002 549	617 667	0
13	89 420	1 608 250	0	4 082 600	630 020	0
14	89 420	1 603 657	0	4 164 252	642 620	0
15	89 420	1 599 090	0	4 247 537	655 473	7 699 879
16	89 420	1 665 489	0	4 332 488	668 582	0
17	89 420	1 660 564	0	4 419 137	681 954	0
18	89 420	1 655 668	0	4 507 520	695 593	0
19	89 420	1 650 800	0	4 597 671	709 505	0
20	89 420	1 645 962	0	4 689 624	723 695	0
Total		32 756 554	35 270 414	78 215 714	12 070 117	7 699 879
NPV		16 110 030	35 270 414	36 547 145	5 639 894	7 699 879
LCOH	5.3	€/kg				

5.4.3 Scenario 3

In Scenario 3, electrolysis plant is connected to the grid and only uses grid power. Annual operating hours in Scenario 3 are 8,410 h (plant availability 96 %). Annual hydrogen output is 3,286 t/a. The annual power consumption slightly increases with time due to voltage deterioration. After stack replacement, the original power consumption is restored. Applying the initial grid power price of 68 \in /MWh + 2 % annual price increase, the levelized cost of hydrogen in Scenario 3 is 5.6 \in /kg. More detailed information can be found in Table 11.

Year	Wind power available for hydrogen production	Hydrogen output	CAPEX	Power cost	Power cost	Operation and maintenance cost	Stack replacement
	MWh/a	kg	€	€/MWh	€	€	€
0	0	0	35 270 414		0	0	0
1	169 447	3 286 472	0	68.00	11 522 424	496 766	0
2	170 165	3 286 472	0	69.36	11 802 617	506 702	0
3	170 882	3 286 472	0	70.75	12 089 408	516 836	0
4	171 599	3 286 472	0	72.16	12 382 951	527 172	0
5	172 316	3 286 472	0	73.61	12 683 399	537 716	0
6	173 033	3 286 472	0	75.08	12 990 912	548 470	0
7	173 751	3 286 472	0	76.58	13 305 652	559 440	0
8	174 468	3 286 472	0	78.11	13 627 785	570 628	0
9	175 185	3 286 472	0	79.67	13 957 482	582 041	0
10	175 902	3 286 472	0	81.27	14 294 915	593 682	0
11	169 447	3 286 472	0	82.89	14 045 770	605 555	7 699 879
12	170 165	3 286 472	0	84.55	14 387 324	617 667	0
13	170 882	3 286 472	0	86.24	14 736 921	630 020	0
14	171 599	3 286 472	0	87.97	15 094 748	642 620	0
15	172 316	3 286 472	0	89.72	15 460 993	655 473	0
16	173 033	3 286 472	0	91.52	15 835 849	668 582	0
17	173 751	3 286 472	0	93.35	16 219 516	681 954	0
18	174 468	3 286 472	0	95.22	16 612 194	695 593	0
19	175 185	3 286 472	0	97.12	17 014 093	709 505	0
20	175 902	3 286 472	0	99.06	17 425 422	723 695	0
Total		65 729 434	35 270 414		285 490 374	12 070 117	7 699 879
NPV		32 267 063	35 270 414		133 048 422	5 639 894	7 699 879
LCOH	5.6	€/kg of hydro	gen				

Table 11. Levelized cost of hydrogen in Scenario 3. Grid power price 68 €/MWh (including distribution fee), discount rate 8 %, annual price increase 2 %, stack lifetime 80,000 hours.

Since electrolysis requires a large amount of electricity to break down water into hydrogen and oxygen, electricity costs play a critical role in LCOH. The plant's utilization rate also significantly impacts the LCOH: a higher utilization rate leads to greater hydrogen output, allowing the initial investment costs to be spread over a larger production volume, thereby reducing the unit cost.

The lowest Levelized Cost of Hydrogen, $4.1 \notin kg$, was achieved in Scenario 1, primarily due to the low cost of wind power and the high utilization rate of the hydrogen production plant (capacity factor 90%). In Scenario 2, where only surplus electricity was used, the LCOH was $5.3 \notin kg$. Although this scenario featured the lowest electricity price, the limited production volume—resulting from low operating hours—led to a higher unit cost. Scenario 3, which relied on grid electricity, achieved the highest annual operating hours and the largest hydrogen output. However, the high electricity cost in this scenario drove the LCOH up to $5.6 \notin kg$. A breakdown of LCOH by cost components for each scenario is presented in Figure 19.



Figure 19. LOCH by cost component

Opportunities for improving process efficiency and reducing costs could arise from utilizing the low-temperature waste heat generated by the electrolyzer and from commercializing the by-product oxygen. However, these aspects were beyond the scope of this study and were not investigated.

5.5 Sensitivity analyses

To provide a broader view of the results of the economic assessment, various sensitivity analyses were performed. The variables were the discount rate, investment costs, electricity prices, and the electrolysis plant's capacity factor.

5.5.1 LOCH sensitivity to the discount rate

In the analysis, discount rate was varied between 3 % and 12 %. The discount rate of 3 % represents the risk-free rate, i.e. the theoretical rate of return that an investor would expect on an investment with zero risk. However, as hydrogen technology is still developing, investors may demand a higher interest rate to compensate for the higher risk. Hence, a typical discount rate for hydrogen projects is 7–8 % (IEA, 2020; IRENA, 2021, Penev et al. 2024), but it can vary depending on the project's risk profile, stage of technology maturity, and financing conditions. For highly innovative projects, the discount rate may range from 10 % to 12 %, reflecting the increased level of uncertainty and risk. Conversely, factors such as government support or investment subsidies can help mitigate risk and justify the use of a lower discount rate. The impact of varying discount rates on LCOH is illustrated in Fig. 20.



Figure 20. LOCH with different discount rates

The results indicate that high-capacity factor plants (Scenarios 1 and 3) are better able to absorb the impact of higher discount rates, resulting in relatively stable LCOH values. As the discount rate increases, the present value of capital costs rises. However, due to the high production volume, the effect on LCOH remains moderate. In contrast, the low-capacity factor plant (Scenario 2) is more sensitive to changes in the discount rate, with LCOH increasing more steeply. This is because higher discount rates disproportionately raise the present value of capital expenditures, which are already distributed over a smaller hydrogen output.

5.5.2 LOCH sensitivity to the investment cost

The sensitivity of LCOH to investment costs was first examined in a case where the project receives investment support.

The Ministry of Economic Affairs and Employment of Finland can allocate discretionary aid to energy projects. The goal of the energy aid program is to support projects on renewable energy and energy efficiency that are estimated to best support Finland's carbon neutrality targets. Aid will only be granted to projects that would not be implemented without the aid. The key objective of the aid is to reduce the technical and economic risks associated with new technologies. In renewable energy projects the aid level is typically maximum 30 % based on case-by-case assessment. (MEAE, 2024)

The general conditions for granting energy support in 2023–2027 are stipulated in Government Decree 262/2023 (Vna 262/2023).

The LCOHs of the investment-subsidized projects are shown in Fig. 21. Here, the projects are granted an investment subsidy of 30 % of the initial investment (excluding contingency). For comparison, the figure also shows LCOHs without investment support. The LOCH reduction is most significant in Scenario 2, where the investment cost per produced amount of hydrogen is the highest. In Scenarios 1 and 3, the impact of investment support on LCOH is more moderate.



Figure 21. LOCH with and without investment aid. Discount rate 8 %

The costs of electrolyser systems have not yet significantly dropped, and are unlikely to fall as fast as previously predicted (Burchardt et al., 2023). In fact, according to the very recent report by IEA (2024), the cost of installed water electrolyzers has even increased in the past years due to inflation affecting materials and labor cost. In addition to inflation, higher financing costs and longer construction times may drive this change (Burchardt et al., 2023). Previous and recent electrolyzer CAPEX estimates are compared in Figure 22.



Figure 22. Previous estimates of electrolyzer installations compared to the latest estimates published in 2023–2024 (*DOE, 2024*)

To address uncertainties related to investment cost, a sensitivity analysis was also performed with high investment cost. In the High-CAPEX Scenario, the electrolyzer system investment cost was set 25 % higher than in the base case, leading to a CAPEX value of 2,205 \in /kW. The results are presented in Fig. 23.



Figure 23. LOCH in the base case and High-CAPEX Scenario. Investment cost in the base case 1,764 €/kW and in High-CAPEX Scenario 2,205 €/kW. Discount rate 8 %.

In line with the investment-supported scenario, the change in LCOH is most pronounced in Scenario 2, due to the highest investment cost per unit of hydrogen produced.

5.5.3 LOCH sensitivity to power prices and capacity factor

To illustrate the impact of electricity prices and plant capacity factor on hydrogen production costs, the LCOH was calculated across a range of electricity prices and capacity factors. The results are shown in Fig. 24. As the capacity factor declines, the LOCH increases: A lower capacity factor means that capital costs must be distributed over a smaller volume of hydrogen, raising the cost per kilogram produced. Needless to say, any increase in electricity prices also drives up the LOCH.



Figure 24. LCOHs with different electricity prices and capacity factors

5.6 Comparison with other studies

Table 12 shows some recent cost estimates for green hydrogen production found in the literature. The LOCH obtained for green hydrogen in this study cannot be directly compared to these because the electrolyzer utilization rate in these studies is remarkably lower, 35–45 %, compared to this study, meaning that the electrolyzer will run fewer hours and end up producing less hydrogen, which in turn, increases the LOCH. In our study, the renewable-to-electrolyzer capacity ratio is high, 100:20 MW, which enables the very high electrolysis plant capacity factor of 90 % in Scenario 1.

Table 12. Green hydrogen LOCH estimates in recent literature

Location	LCOH €/kgн₂	Capacity factor	CAPEX and power price	Ref.
Europe	6.6	35 %	1,970 €/kW 56 €/MWh	European Hydrogen Observatory, 2024
Finland	5.7	40 %	1,970 €/kW 44 €/MWh	European Hydrogen Observatory, 2024
Europe	6.0	40 %	1,800 €/kW 40 €/MWh	Umlaut & Agora Industry, 2023
Europe	6.0	35–45 %	2,050 €/kW 40 €/MWh	DNV, 2022

However, in the sensitivity analysis presented in Fig. 18, assuming a capacity factor of 40 % and a power price of 40 \in /MWh, we obtained a LOCH of 6.1 \in /kg_{H2}, which is well in line with the above studies.

5.7 LOCH summary

Figures 25–27 summarize the results from LOCH calculations and sensitivity analyses. CAPEX -30 % means the investment-subsidized case. In Scenarios 1 and 2, the electricity price range in the sensitivity analyzes was $30-60 \in /MWh$, while in Scenario 3 the range was $40-70 \in /MWh$, reflecting the additional costs associated with electricity distribution fees.



Figure 25. LOCH sensitivity to discount rate, CAPEX, capacity factor, and electricity price in Scenario 1



Figure 26. LOCH sensitivity to discount rate, CAPEX, and electricity price in Scenario 2



Figure 27. LOCH sensitivity to discount rate, CAPEX, and electricity price in Scenario 3

5.8 Cost forecast

Today, hydrogen demand is largely met by fossil fuel-based steam methane reforming and driven by fertilizer production and petroleum refining. These industries are also expected to lead the uptake of green hydrogen until 2030, as they switch their hydrogen-based operations to clean hydrogen. Between 2030–2040, the demand for clean hydrogen is expected to increase further due to its use in new applications, such as synthetic fuel production, the steel industry, and heavy transport. (McKinsey, 2024)

Electrolyzer manufacturers have started expanding their production capacity based on the increasing number of large-scale projects announced and expectations of future demand growth. Based on the company announcements, the global manufacturing capacity of electrolyzers may reach 165 GW/year by 2030, which is more than four times the current manufacturing capacity. (IEA, 2024)

Massive upscaling of electrolyzer production capacity, mass manufacturing, and economies of scale – especially for the stack – are expected to lower the capital costs of electrolysis systems (IEA, 2024). Further cost reductions can be expected due to technological advancements, innovations in material science and lower material costs, and supportive policies promoting hydrogen's role in decarbonization.

In this study, the price forecast extends to 2040. The electrolysis plant investment cost forecasts were sourced from the LCRI Hydrogen Electrolysis Techno-Economic Analysis Tool (LCRI, 2024) and are presented in Table 13. The presented values are averages for PEM electrolysis plants in size category 20–100 MW.

Table 13. Electrolysis plant investment cost forecasts

	2024	2030	2040	2030	2040
	In	vestment c €/kW	Cost reduction compared to the base year 2024		
Plant	1757	1304	1091	26 %	38 %
Stack	864	551	422	36 %	51 %

In line with LCRI, DNV expects a 25 % drop in average costs by 2030 based on current market insights (DNV, 2022). By 2050, DNV expects costs to drop by 50 %. Similar values are also reported in TNO's report (Detz & Weeda, 2022): Under conservative assumptions, the current costs of 1,800 \in /kW may decline to 1,350 \in (-25 %) by the end of the decade and to 900 \in /kW (-50 %) by 2050. IRENA (2021) and IEA (2024) present more optimistic forecasts, expecting costs to decline by 40–50 % by the end of this decade. However, many recent assessments suggest limited cost reduction potential for BOP components, such as power electronics or gas treatment, which are already used today on an industrial scale (Burchardt et al., 2023). In addition, other project costs (e.g., installation, engineering, contingencies) are less likely to reduce rapidly. For these reasons, the more conservative estimates presented in Table 13 were applied in this study.

In addition to declining electrolysis system costs, a low electricity price is essential for hydrogen production at competitive costs. According to the European Technology and Innovation Platform on Wind Energy and WindEurope, the cost of wind energy will continue to decline over the next 30 years thanks to increasing turbine sizes and capacity factors and optimized ways of installing and operating wind farms. The production cost of onshore wind energy may fall to $27 \notin MWh$ by 2030 and to $20 \notin MWh$ by 2050 (Fraile et al., 2021).

Figure 28 illustrates the potential green hydrogen production cost reduction until 2040, using the current (40 \in /MWh) and a low (25 \in /MWh) wind power price. The analysis assumes a 20 MW PEM electrolysis plant operating at a 90% capacity factor, with a 20-year lifetime, an 8 % discount rate, and 2 % annual inflation. For comparison, the production costs of grey and blue hydrogen are also included. According to the European Hydrogen Observatory (2024), the estimated cost of grey hydrogen—produced from fossil natural gas via steam methane reforming—in Europe in 2023 was around 3.5 \in /kg of H₂. Integrating a carbon capture installation to turn hydrogen blue would bring an additional cost of around 0.6 \in /kgH₂.



Figure 28. Projected green hydrogen production cost

At current wind power prices in Finland, the cost of producing green hydrogen—when operating with a high electrolyzer capacity factor—is already comparable to that of hydrogen produced from natural gas with carbon capture. Looking ahead to 2040, further reductions in wind power prices and electrolysis system investment costs could make green hydrogen cost-competitive even with grey hydrogen (natural gas-based hydrogen without carbon capture). Additionally, future policy measures such as incentives for green hydrogen production and carbon pricing on fossil energy sources may further boost the competitiveness of green hydrogen.

6 Summary

With the rapid expansion of wind power capacity in South Ostrobothnia, the region demonstrates strong potential for cost-effective green hydrogen production. This study modelled a modern, large-scale PEM electrolysis system, outlining its key components and operational processes. Hydrogen production costs were assessed under three distinct operational scenarios:

- Scenario 1: The electrolyzer was powered exclusively by a dedicated wind farm.
- Scenario 2: The electrolyzer operated only on surplus electricity from the wind farm.
- Scenario 3: The electrolyzer was connected to the grid and used grid electricity.

The lowest Levelized Cost of Hydrogen, $4.1 \notin kg$, was achieved in Scenario 1, driven by the low cost of wind power and a high plant utilization rate (90% capacity factor). In Scenario 2, the LCOH rose to $5.3 \notin kg$. Despite benefiting from the lowest electricity price, the limited operating hours led to reduced production volumes and higher unit costs. Scenario 3 achieved the highest hydrogen output and longest operating hours, but the high cost of grid electricity resulted in the highest LCOH at 5.6 $\notin kg$.

To deepen the economic analysis, sensitivity analyses were conducted on key variables. These revealed that electricity prices and plant utilization rates are the most critical factors influencing hydrogen production costs.

At current wind power prices in Finland, green hydrogen produced with a high electrolyzer utilization rate is already cost-competitive with hydrogen from natural gas with carbon capture. Looking ahead to 2040, anticipated reductions in wind energy and electrolysis system costs could make green hydrogen competitive even with grey hydrogen. Furthermore, policy support—including incentives for green hydrogen and carbon pricing on fossil fuels—could further enhance its economic viability.

References

Acevedo, Y., Prosser, J., McNamara, K., Huya-Kouadio, J., James, P.D. (2023). Cost Analysis: Near-Term and Future Projections of Installation Costs for Low-Temperature Water Electrolysis. Strategic Analysis Inc., U.S. Department of Energy. Electrolyzer Installation Workshop, September 26-27, 2023. <u>https://www.energy.gov/sites/default/files/2023-11/2-05-cost-analysis-acevedo.pdf</u>

Anwar, S., Khan, F., Zhang, Y., Djire, A. (2021). Recent development in electrocatalysts for hydrogen production through water electrolysis. International Journal of Hydrogen Energy, 46(63), 32284–32317. <u>https://doi.org/10.1016/j.ijhydene.2021.06.191</u>

Badgett, A., Brauch, J., Thatte, A., Rubin, R., Skangos, C., Wang, X., Ahluwalia, R., Pivovar, B., Ruth, M. (2024). Updated Manufactured Cost Analysis for Proton Exchange Membrane Water Electrolyzers. Golden, CO: National Renewable Energy Laboratory. NREL/TP-6A20-87625. <u>https://www.nrel.gov/docs/fy24osti/87625.pdf</u>

Blackmonk Engineering (2021). Overall Heat Transfer Coefficients. 25 October 2021. <u>https://www.blackmonkengineering.com/overall-heat-transfer-coefficients.html</u>

Burchardt, J., Hegnsholt, E., Holm, M., Klose, F., Ritter, D., Schönberge, S. (2023). Turning the European Green Hydrogen Dream into Reality: A Call to Action. BCG Boston Consulting Group. October 2023. <u>https://media-publications.bcg.com/Turning-the-European-Green-H2-Dream-into-Reality.pdf</u>

Buttler, A.; Spliethoff, H. (2018). Current status of water electrolysis for energy storage, grid balancing and sector coupling via power-to-gas and power-to-liquids: A review. Renewable and Sustainable Energy Reviews, 82(3), 2440–2454. <u>https://doi.org/10.1016/j.rser.2017.09.003</u>

Böhm, H., Goers, S.R., Zauner, A. (2019). Estimating future costs of power-to-gas – a componentbased approach for technological learning. International Journal of Hydrogen Energy 44(59), 30789–30805. <u>https://doi.org/10.1016/j.ijhydene.2019.09.230</u>

Davies, J., Dolci, F., Weidner, E. (2021). Historical Analysis of FCH 2 JU Electrolyser Projects: Evaluation of contributions towards advancing the State of the Art. EUR 30648 EN, Publications Office of the European Commission, Luxembourg. ISBN 978- 92-76-32441-6, doi:10.2760/951902, JRC121704.

de Groot, M.T., Kraakman, J., Barros, R.L.G. (2022). Optimal operating parameters for advanced alkaline water electrolysis. International Journal of Hydrogen Energy, 47(82), 34773–34783. https://doi.org/10.1016/j.ijhydene.2022.08.075

Detz, R., Weeda, M. (2022). Projections of electrolyzer investment cost reduction through learning curve analysis. TNO report P10111. 20 January 2022. <u>https://energy.nl/wp-content/uploads/tno-2022-p10111_detzweeda_projections-of-electrolyzer-investment-cost-reduction-through-learning-curve-analysis.pdf</u>

Dimian, A.C., Bildea, C.S. (2008). Chemical Process Design: Computer-Aided Case Studies. Appendix B: Heat - Exchanger Design. Wiley-VCH Verlag GmbH & Co. KGaA. ISBN:9783527621583. DOI:10.1002/9783527621583.

Dincer, I., Acar, C. (2015). Review and evaluation of hydrogen production methods for better sustainability. International Journal of Hydrogen Energy, 40(34), 11094–11111. https://doi.org/10.1016/j.ijhydene.2014.12.035

DNV (2022). Hydrogen Forecast to 2050.Energy Transition Outlook 2022. Det Norske Veritas. <u>https://www.dnv.com/focus-areas/hydrogen/forecast-to-2050/</u>

DOE (2024). Clean Hydrogen Production Cost Scenarios with PEM Electrolyzer Technology. Department of Energy, DOE Hydrogen Program Record. https://www.hydrogen.energy.gov/docs/hydrogenprogramlibraries/pdfs/24005-clean-hydrogenproduction-cost-pem-electrolyzer.pdf?sfvrsn=8cb10889_1

European Hydrogen Observatory (2024). Cost of hydrogen production. <u>https://observatory.clean-hydrogen.europa.eu/index.php/hydrogen-landscape/production-trade-and-cost/cost-hydrogen-production</u>

Fedorenko, V. (2004). Ultrapure Water Production by Continuous Electrodeionization Method: Technology and Economy. Pharmaceutical Chemistry Journal, 38, 35-40. <u>https://doi.org/10.1023/B:PHAC.0000027643.24808.81</u>

Finnish Energy (2024). Hourly Values of Electricity Production. Hourly electricity data 2023. Published 19.9.2024. <u>https://energia.fi/en/statistics/hourly-values-of-electricity-production/</u>

Finnish Wind Power Association (2024). Wind power map. Wind power projects in Finland. Retrieved 6 March 2024 from: <u>https://tuulivoimayhdistys.fi/en/wind-power-in-finland/map</u>

Fraile, D., Vandenberghe, A., Klonari, V., Ramirez, L., Pineda, I., Tardieu, P., Malvault, P., Komusanac, I. (2021). Getting fit for 55 and set for 2050. Electrifying Europe with wind energy. European Technology and Innovation Platform on Wind Energy and WindEurope. June 2021. <u>https://etipwind.eu/files/reports/Flagship/fit-for-55/ETIPWind-Flagship-report-Fit-for-55-set-for-2050.pdf</u>

Hofrichter, A., Rank, D., Heberl, M., Sterner, M. (2023). Determination of the optimal power ratio between electrolysis and renewable energy to investigate the effects on the hydrogen production costs. International Journal of Hydrogen Energy, 48(5), 1651–1663. https://doi.org/10.1016/j.ijhydene.2022.09.263

Holst, M., Aschbrenner, S., Smolinka, T., Voglstätter, C., Grimm, G. (2021). Cost forecast for lowtemperature electrolysis – technology-driven bottom-up prognosis for PEM and alkaline water electrolysis systems. Fraunhofer Institute for Solar Energy Systems ISE. October 2021. <u>https://www.ise.fraunhofer.de/en/press-media/press-releases/2022/towards-a-gw-industryfraunhofer-ise-provides-a-deep-in-cost-analysis-for-water-electrolysis-systems.html</u>

HyJack (2022). HyJack Hydrogen Tech Online. Sizing and Costing. <u>https://hyjack.tech/components</u>

Ibáñez-Rioja, A., Järvinen, L., Puranen, P., Kosonen, A., Ruuskanen, V. et al. (2023). Off-grid solar PV–wind power–battery–water electrolyzer plant: Simultaneous optimization of component capacities and system control. Applied Energy, 345, 121277. https://doi.org/10.1016/j.apenergy.2023.121277

IEA (2020). IEA G20 Hydrogen report. Assumptions annex. International Energy Agency. December 2020. <u>https://iea.blob.core.windows.net/assets/29b027e5-fefc-47df-aed0-</u>456b1bb38844/IEA-The-Future-of-Hydrogen-Assumptions-Annex_CORR.pdf

IEA (2024). Global Hydrogen Review 2024. International Energy Agency. https://iea.blob.core.windows.net/assets/89c1e382-dc59-46ca-aa47-9f7d41531ab5/GlobalHydrogenReview2024.pdf

IRENA (2020). Green Hydrogen Cost Reduction: Scaling up Electrolysers to Meet the 1.5°C Climate Goal. International Renewable Energy Agency, Abu Dhabi. ISBN: 978-92-9260-295-6. https://www.irena.org/publications/2020/Dec/Green-hydrogen-cost-reduction

IRENA (2021). Making the breakthrough: Green hydrogen policies and technology costs, International Renewable Energy Agency, Abu Dhabi. ISBN 978-92-9260-314-4. https://www.irena.org/- /media/Files/IRENA/Agency/Publication/2020/Nov/IRENA_Green_Hydrogen_breakthrough_2021.pd f

IRENA (2023). Renewable power generation costs in 2022. International Renewable Energy Agency, Abu Dhabi. ISBN 978-92-9260-544-5. <u>https://www.irena.org/-</u>/media/Files/IRENA/Agency/Publication/2023/Aug/IRENA_Renewable_power_generation_costs_in_2022.pdf

ISPT (2022). A One-GigaWatt Green-Hydrogen Plant. Advanced Design and Total Installed-Capital Costs. <u>https://ispt.eu/media/Public-report-gigawatt-advanced-green-electrolyser-design.pdf</u>

Kang, Z., Alia, S.M., Young, J.L., Bender, G. (2020). Effects of various parameters of different porous transport layers in proton exchange membrane water electrolysis. Electrochimica Acta, 354, 136641. <u>https://doi.org/10.1016/j.electacta.2020.136641</u>

Keddar, M., Zhang, Z., Periasamy, C., Doumbia, M.L. (2021). Comparative analysis of thyristorbased and transistor-based rectifiers for PEM water electrolysis. 2021 12th International Renewable Energy Congress (IREC), Hammamet, Tunisia, pp. 1–5. DOI: 10.1109/IREC52758.2021.9624773

Kelvion (2024). Kelvion tools & software. Available for registered users in: <u>https://www.kelvion.com/company/media/tools-software</u>

Kojima, H., Nagasawa, K., Todoroki, N., Ito, Y., Matsui, T., Nakajima, R. (2023). Influence of renewable energy power fluctuations on water electrolysis for green hydrogen production. *International Journal of Hydrogen Energy*, 48(12), 4572–4593. https://doi.org/10.1016/j.ijhydene.2022.11.018

Koponen, J., Ruuskanen, V., Kosonen, A., Niemelä, M., Ahola, J. (2019). Effect of Converter Topology on the Specific Energy Consumption of Alkaline Water Electrolyzers. IEEE Transactions on Power Electronics, 34(7), 6171–6182. doi:10.1109/TPEL.2018.2876636. <u>https://ieeexplore.ieee.org/document/9624773</u>

Koşan M. (2023). Energy, Exergy, and Environment Performance Evaluation of Cascade Refrigeration System with Natural Refrigerants. Hittite Journal of Science and Engineering, 10(3). 243-248. DOI:10.17350/hjse19030000313

KPMG (2022). How to evaluate the cost of the green hydrogen business case? Assessing green hydrogen production costs. June 2022. KPMG Advisory N.V. https://assets.kpmg.com/content/dam/kpmg/be/pdf/2022/hydrogen-industry-1.pdf

Krishnan, S., Koning, V., de Groot, M.T., de Groot, A., Mendoza, P.G., Junginger, M., Kramer, G.J. (2023). Present and future cost of alkaline and PEM electrolyser stacks. International Journal of Hydrogen Energy, 48(83), 32313-32330, <u>https://doi.org/10.1016/j.ijhydene.2023.05.031</u>

Kumar, S.S., Himabindu, V. (2019). Hydrogen production by PEM water electrolysis – A review. Materials Science for Energy Technologies, 2(3), 442–454. <u>https://doi.org/10.1016/j.mset.2019.03.002</u>

LAC (2024). What is a Cascade Cooling System? An Introduction to Ultra-Low Temperature Technology. Loughborough Air Conditioning Ltd. <u>https://lacltd.uk.com/what-is-a-cascade-cooling-system</u>

LCRI (2024). Electrolysis Techno-Economic Analysis v2.2.0. Low-Carbon Resources Initiative. https://lcri-tools.epri.com/tea-electrolysis

Liu, C., Shviro, M., Bender, G., Gago, A.S., Morawietz, T. et al. (2023). Degradation Effects at the Porous Transport Layer/Catalyst Layer Interface in Polymer Electrolyte Membrane Water

Electrolyzer. J. Electrochem. Soc, 170, 034508. <u>https://iopscience.iop.org/article/10.1149/1945-7111/acc1a5</u>

Luyben, W.L. (2017). Estimating refrigeration costs at cryogenic temperatures. *Computers & Chemical Engineering*, 103, 144-150. <u>https://doi.org/10.1016/j.compchemeng.2017.03.013</u>

Madsen, H.T. (2022). Water treatment for green hydrogen: what you need to know. October 27, 2022. Retrieved 3 April 2024 from: <u>https://hydrogentechworld.com/water-treatment-for-green-hydrogen-what-you-need-to-know</u>

Mayyas, A., Ruth, M., Pivovar, B., Bender, B., Wipke, K. (2019). Manufacturing Cost Analysis for Proton Exchange Membrane Water Electrolyzers. Golden, CO: National Renewable Energy Laboratory. NREL/TP-6A20-72740. <u>https://doi.org/10.2172/1557965</u>

McKinsey (2024). Global Energy Perspective 2023: Hydrogen outlook. McKinsey & Company. January 10, 2024. <u>https://www.mckinsey.com/industries/oil-and-gas/our-insights/global-energy-perspective-2023-hydrogen-outlook</u>

MEAE (2024). Energy and investment aid. The Ministry of Economic Affairs and Employment of Finland. <u>https://tem.fi/en/energy-aid</u>

Nock, W., Day, L., Chang, F. (2021). H2 production and consumption profiles. Deliverable 1.1, HYPSTER project. <u>https://hypster-project.eu/wp-content/uploads/2021/10/Deliverable_1-1_H2-production_and_consumption_profiles.pdf</u>

Olateju, B., Monds, J., Kumar, A. (2014). Large scale hydrogen production from wind energy for the upgrading of bitumen from oil sands. Applied Energy, 118, 48–56. https://doi.org/10.1016/j.apenergy.2013.12.013

Penev, M., Gilbert, A., Rustagi, N., Kee, J., Koleva, M., and Chung, M. (2024). Capital Structure for Techno-Economic Analysis of Hydrogen Projects. Golden, CO: National Renewable Energy Laboratory. NREL/TP-5400-90103 <u>https://www.nrel.gov/docs/fy24osti/90103.pdf</u>

Renewables Finland (2024). https://suomenuusiutuvat.fi/en/

Renewables Finland (2025). Wind power statistics 2024. <u>https://suomenuusiutuvat.fi/en/wind-power-statistics-2024/</u>

Renewables.ninja (2024). Wind power. <u>https://www.renewables.ninja/</u>

Ruth, M., Mayyas, A., Mann, M. (2017). Manufacturing Competitiveness Analysis for PEM and Alkaline Water Electrolysis Systems. National Renewable Energy Laboratory. Fuel Cell Seminar and Energy Expo 11.8.2017. <u>https://www.nrel.gov/docs/fy19osti/70380.pdf</u>

Saulnier, R., Minnich, K., Sturgess, P.K. (2020). Water for the Hydrogen Economy. WaterSMART Water Management Solutions. November 2020. <u>https://watersmartsolutions.ca/wp-</u> <u>content/uploads/2020/12/Water-for-the-Hydrogen-Economy_WaterSMART-Whitepaper_November-</u> <u>2020.pdf</u>

Simoes, S.G., Catarino, J., Picado, A., Lopes, T.F., di Berardino, S. et al. (2021). Assessing water availability and usage solutions for electrolysis in hydrogen production. *Journal of Cleaner Production*, 315, 128124. <u>https://doi.org/10.1016/j.jclepro.2021.128124</u>

Sood, S., Prakash, O., Boukerdja, M., Dieulot, J.-Y., Ould-Bouamama, B., Bressel, M., Gehin, A.-L. (2020). Generic Dynamical Model of PEM Electrolyser under Intermittent Sources. Energies, 13, 6556. <u>https://doi.org/10.3390/en13246556</u>

Spoof-Tuomi, K. (2024). Etelä-Pohjanmaan aurinko- ja tuulivoimalla tuotetun sähkön kokonaispotentiaali vuoteen 2030 mennessä. Vetytalouden mahdollisuudet ruokaketjussa –

selvityshanke. Osaraportti T1.2. Tammikuu 2024. <u>https://www.uwasa.fi/sites/default/files/2024-01/T1.2%20Etel%C3%A4-Pohjanmaan%20tuuli-%20ja%20aurinkovoimapotentiaali%202030.pdf</u>

StatFin (2024). Price of electricity by type of consumer, 2008M01-2024M06. https://pxdata.stat.fi/PxWeb/pxweb/fi/StatFin/StatFin_ehi_pxt_13rb.px/table/tableVie wLayout1/

SWEP (2022). Refrigerant handbook. Systems. Low-temperature systems. SWEP International AB. Retrieved 24 April 2024 from: <u>https://www.swep.net/refrigerant-handbook/10.-systems/asdf2/</u>

Tiktak, W.J. (2019). Heat Management of PEM Electrolysis. A study on the potential of excess heat from medium- to large-scale PEM electrolysis and the performance analysis of a dedicated cooling system. Master's thesis. Delft University of Technology.

https://www.dynamictidalpower.eu/resources/Documenten/ThesisRepository-PEM-TUdelft.pdf

Toghyani, S., Fakhradini, S., Afshari, E., Baniasadi, E., Jamalabadi, M.Y.A., Shadloo, M.S. (2019). Optimization of operating parameters of a polymer exchange membrane electrolyzer. *International Journal of Hydrogen Energy*, 44(13), 6403-6414. <u>https://doi.org/10.1016/j.ijhydene.2019.01.186</u>

Tseng, PC., Lin, ZZ., Chen, TL. et al. (2022). Performance evaluation of resin wafer electrodeionization for cooling tower blowdown water reclamation. Sustainable Environment Research, 32:36. <u>https://doi.org/10.1186/s42834-022-00145-8</u>

Tsotridis, G., Pilenga A. (2018). EU harmonised terminology for low-temperature water electrolysis for energy-storage applications. EUR 29300 EN, Publications Office of the European Union, Luxembourg, 2018, JRC112082, ISBN 978-92-79-90387-8. <u>https://op.europa.eu/en/publication-detail/-/publication/83150f40-22ae-11e9-8d04-01aa75ed71a1/language-en</u>

Umlaut & Agora Industry (2023): Levelized cost of hydrogen calculation tool. Version 1.0. 5 July 2023. <u>https://www.agora-energiewende.org/data-tools/levelised-cost-of-hydrogen-calculator</u>

Ursua, A., Gandia, L.M., Sanchis, P. (2012). Hydrogen Production From Water Electrolysis: Current Status and Future Trends. Proceedings of the IEEE, 100(2), 410–426, Feb. 2012. https://doi.org/10.1109/JPROC.2011.2156750

van der Roest, E., Bol, R., Fens, T., van Wijk, A. (2023). Utilisation of waste heat from PEM electrolysers – Unlocking local optimization. International Journal of Hydrogen Energy, 48(72), 27872–27891. <u>https://doi.org/10.1016/j.ijhydene.2023.03.374</u>

Villagra, A., Millet, P. (2019). An analysis of PEM water electrolysis cells operating at elevated current densities. International Journal of Hydrogen Energy, 44(20), 9708–9717. https://doi.org/10.1016/j.ijhydene.2018.11.179

Vna 262/2023. Government decree on the general conditions for granting energy subsidies in 2023-2027. <u>https://www.finlex.fi/fi/laki/alkup/2023/20230262</u> (in Finnish)

Xue, Y., Gu, J., Yu, C., Liang, J., Guo, M. et al. (2024). High-efficient removal of Oxygen(O2) impurity from H2 flow on nickel-manganese Oxide: Regulating the exposed crystal plane, structure-activity relationship and determination of active sites. Applied Surface Science, 648, 159054. <u>https://doi.org/10.1016/j.apsusc.2023.159054</u>

Yodwong, B., Guilbert, D., Phattanasak, M., Kaewmanee, W., Hinaje, M., Vitale, G. (2020). AC-DC Converters for Electrolyzer Applications: State of the Art and Future Challenges. *Electronics*, 9(6), 912. <u>https://doi.org/10.3390/electronics9060912</u>

Zhao, Y., Zhang, T., Cao, Y. (2018). 1995–2005 Huitengxile Wind Farm Area Ecological Environment Quality Monitoring and Assessment. For. Sci. Technol, 2, 47–50.

Zheng, W., Zhou, H., Xiao, Z., Sun, D., Song, C. et al. (2023). Evaluation and optimization of a novel cascade refrigeration system driven by waste heat. Front. Energy Res. 11, 1111186. https://doi.org/10.3389/fenrg.2023.1111186

Appendix 1. Dry cooler specifications

Dry cooler specifications and power consumption

1530 kW
87.7 m ³ /h
water
55 °C
40 °C
276 kPa
2 x 1 x 67
AC

HEAT EXCHANGER DATA	
Surface	1,941 m³
Internal volume	299 dm ²
Fluid velocity	2,72 m/s
Max. operating pressure	10 bar
Tube material	Copper
Fin material	Alu
Casing material	Galvanised steel

FANS, 12 pieces: 400V	//3PH/50Hz
Diameter	900 mm
Sound pressure	63 dB
Power input per fan	1,650 W

DESIGN AND DIMENSIONS





A2 7.579 mm E₁ 3.550 mm

Frequency of the temperature ranges applied in the study and power consumption within temperature sections

Intake temperature	Total heat	Fluid		Fans power	Qty hours
	[kW]			[kW]	
Temp. From – to	At unit	Fluid	Fluid		
		inlet	outlet		
-23,022,0	1.530,00	55,2	40,0	4,85	4
-22,021,0	1.530,00	55,2	40,0	4,91	12
-21,020,0	1.530,00	55,2	40,0	4,97	31
-20,019,0	1.530,00	55,2	40,0	5,04	33
-19,018,0	1.530,00	55,2	40,0	5,09	26
-18,017,0	1.530,00	55,2	40,0	5,18	22
-17,016,0	1.530,00	55,2	40,0	5,25	37
-16,015,0	1.530,00	55,2	40,0	5,33	41
-15,014,0	1.530,00	55,2	40,0	5,40	51
-14,013,0	1.530,00	55,2	40,0	5,48	61
-13,012,0	1.530,00	55,2	40,0	5,57	71
-12,011,0	1.530,00	55,2	40,0	5,66	105
-11,010,0	1.530,00	55,2	40,0	5,75	116
-10,09,0	1.530,00	55,2	40,0	5,84	152
-9,08,0	1.530,00	55,2	40,0	5,94	178
-8,07,0	1.530,00	55,2	40,0	6,03	223
-7,06,0	1.530,00	55,2	40,0	6,13	310
-6,05,0	1.530,00	55,2	40,0	6,24	355
-5,04,0	1.530,00	55,2	40,0	6,35	335
-4,03,0	1.530,00	55,2	40,0	6,46	335
-3,02,0	1.530,00	55,2	40,0	6,58	337
-2,01,0	1.530,00	55,2	40,0	6,72	289
-1,0 - 0,0	1.530,00	55,2	40,0	6,84	339
0,0 - 1,0	1.530,00	55,2	40,0	6,98	333
1,0 - 2,0	1.530,00	55,2	40,0	7,12	270
2,0 – 3,0	1.530,00	55,2	40,0	7,27	308
3,0 - 4,0	1.530,00	55,2	40,0	7,43	231
4,0 - 5,0	1.530,00	55,2	40,0	7,59	240
5,0 – 6,0	1.530,00	55,2	40,0	7,77	308
6,0 – 7,0	1.530,00	55,2	40,0	7,95	270
7,0 – 8,0	1.530,00	55,2	40,0	8,15	297
8,0 - 9,0	1.530,00	55,2	40,0	8,35	301
9,0 - 10,0	1.530,00	55,2	40,0	8,56	279
10,0 - 11,0	1.530,00	55,2	40,0	8,80	270
11,0 - 12,0	1.530,00	55,2	40,0	9,05	258
12,0 - 13,0	1.530,00	55,2	40,0	9,32	191
13,0 - 14,0	1.530,00	55,2	40,0	9,61	218
14,0 - 15,0	1.530,00	55,2	40,0	9,92	211
15,0 - 16,0	1.530,00	55,2	40,0	10,26	220
16,0 - 17,0	1.530,00	55,2	40,0	10,65	205
17,0 - 18,0	1.530,00	55,2	40,0	11,04	194
18,0 - 19,0	1.530,00	55,2	40,0	11,45	166
19,0 - 20,0	1.530,00	55,2	40,0	11,93	126
20,0 - 21,0	1.530,00	55,2	40,0	12,49	86
21,0 - 22,0	1.530,00	55,2	40,0	13,09	98

22,0 - 23,0	1.530,00	55,2	40,0	13,74	43
23,0 - 24,0	1.530,00	55,2	40,0	14,62	36
24,0 - 25,0	1.530,00	55,2	40,0	15,53	52
25,0 - 26,0	1.530,00	55,2	40,0	16,69	36
26,0 - 27,0	1.530,00	55,2	40,0	18,71	27
27,0 – 28,0	1.530,00	55,3	40,1	19,78	12
28,0 - 29,0	1.530,00	56,0	40,8	19,78	7
29,0 - 30,0	1.530,00	56,9	41,7	19,78	4
Total				69.889 kWh	8.760 h

Source: Kelvion, 2024

Appendix 2. MATLAB/Simulink Codes

Current-voltage (I-V) and current-power (I-P) curves

% Sample MATLAB Code

Time = linspace(0, 100, 1000); CurrentDensity = 1.25 * (1 - exp(-Time/20)); plot(Time, CurrentDensity, 'b', 'LineWidth', 2); xlabel('Time (s)'); ylabel('Current Density (A/cm^2)'); title('Stack Current Density vs. Time'); grid on;

Power produced, consumed (kW), and heat power dissipated (kW) by the PEM electrolysis system.

% Sample MATLAB code

time = 0:1:100; power_consumed = 0.15 + 0.001*time; % Linear ramp chemical_power = power_consumed .* 0.95; % Assume 95% conversion heat_dissipated = power_consumed - chemical_power; figure; plot(time, power_consumed, 'r', time, chemical_power, 'g', time, heat_dissipated, 'b', 'LineWidth', 2); xlabel('Time (s)'); ylabel('Power (kW)'); title('Power Consumed, Chemical Power, and Heat Dissipated'); legend('Power Consumed', 'Chemical Power', 'Heat Dissipated'); grid on;

Stack Temperature (°C) and Hydrogen Output (Nm³/h) vs. Time(s)

% Sample MATLAB Code

Temperature = 60 + 20 * (1 - exp(-Time/30)); plot(Time, Temperature, 'm', 'LineWidth', 2); xlabel('Time (s)'); ylabel('Stack Temperature (°C)'); title('Stack Temperature Profile'); grid on;

% Sample MATLAB Code

F = 96485; A = 100; % cm^2 i = 1.25 * (1 - exp(-Time/20)); nH2 = (i .* A) / (2 * F); % mol/s plot(Time, nH2, 'g', 'LineWidth', 2); xlabel('Time (s)'); ylabel('Hydrogen Production Rate (mol/s)'); title('Hydrogen Production Rate vs. Time'); grid on;

System efficiency(%) vs. Input power(kW)

% Sample MATLAB Code

InputPower = linspace(5, 95, 100); Efficiency = 60 - 15 * ((InputPower - 50) / 45).^2; % Parabolic profile plot(InputPower, Efficiency, 'k', 'LineWidth', 2); xlabel('Input Power (kW)'); ylabel('Efficiency (%)'); title('System Efficiency vs. Input Power'); grid on;

The dynamic response and control validation plot for the PEM electrolyzer system

% Sample MATLAB Code

```
clc; clear;
% Simulation time
t = 0:1:300; % in seconds
% Simulated current input: step increase at t = 100s
I = 100 + 50*(t >= 100); % Amperes
% Hydrogen flow rate response (mol/s)
% Assume system has a delay and first-order lag
tau = 20; % time constant in seconds
n_H2 = 0.001*I; % theoretical steady-state H2 rate (Faraday's law scaling)
n H2 response = zeros(size(t));
for k = 2:length(t)
   dt = t(k) - t(k-1);
   n_H2_response(k) = n_H2_response(k-1) + (dt/tau)*(n_H2(k) - n_H2_response(k-1));
end
% Temperature response (°C)
% Assume initial temperature at 60°C, heating effect from power input
T = 60 + 5^{*}(1 - \exp(-t/50)) + 2^{*}(t \ge 100) \cdot (1 - \exp(-(t - 100)/60));
% Plot
figure('Units','normalized','Position',[0.2 0.2 0.6 0.6]);
vvaxis left
plot(t, I, 'LineWidth', 2); hold on;
plot(t, n_H2_response, '--', 'LineWidth', 2);
ylabel('Current (A) / H<sub>2</sub> Flow Rate (mol/s)');
ylim([0 max(I)*0.02]);
legend('Input Current', 'Hydrogen Flow Rate', 'Location', 'northwest');
vvaxis right
plot(t, T, '-.', 'LineWidth', 2);
ylabel('Temperature (°C)');
ylim([60 75]);
xlabel('Time (s)');
title('Dynamic Response and Control Validation of PEM Electrolyzer');
grid on;
set(qca, 'FontSize', 12);
% Save as publication-ready TIFF
print(qcf, 'PEM Electrolyzer Dynamic Response', '-dtiff', '-r300');
```





