

Article

Techno-Economic Optimal Operation of an On-Site Hydrogen Refueling Station

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Abstract

An on-site hydrogen refueling station (HRS) directly supplies hydrogen to vehicles using an on-site hydrogen production method such as electrolysis. For the efficient operation of an on-site HRS, it is essential to optimize the entire process from hydrogen production to supply. However, most existing approaches focus on the efficiency of hydrogen production. This study proposes an optimal operation model for a renewable-energy-integrated on-site HRS, which considers the degradation of electrolyzers and operation of compressors. The proposed model maximizes profit by considering the hydrogen revenue, electricity costs, and energy storage system degradation. It estimates hydrogen production using a voltage equation, models compressor power using a shaft power equation, and considers electrolyzer degradation using an empirical voltage model. The effectiveness of the proposed model is evaluated through simulation. Comparison with a conventional control strategy shows an increase of over 56% in the operating revenue.

Keywords: hydrogen refueling station; proton exchange membrane water electrolyzer; hydrogen compressor



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1. Introduction

The adoption of fuel-cell electric vehicles (FCEVs) that utilize hydrogen energy has increased in recent years. By 2030, it is expected that one out of every twelve vehicles sold in various regions, such as Germany, Japan, and California, will be an FCEV [1]. To support this expansion, it is essential to develop infrastructure such as hydrogen refueling stations (HRSs) [2]. In this context, several studies have investigated the techno-economic performance and optimal sizing of HRSs [3,4], and numerous works have reported levelized cost of hydrogen (LCOH) estimates across different station sizes and operating scenarios [5,6].

HRSs can be categorized into off-site and on-site types depending on the hydrogen supply method. An off-site HRS receives externally produced hydrogen via trailers or pipelines. This hydrogen is compressed, stored, and dispensed on site. In this configuration, the hydrogen transport cost accounts for up to 50% of the total supply chain cost [7,8]. In contrast, an on-site HRS directly produces hydrogen at the station using equipment such as electrolyzers or steam methane reforming, followed by compression and storage. This eliminates the requirement for hydrogen transport. However, in the on-site configuration, the electricity consumed for hydrogen production and compression accounts for more than 40% of the total supply chain cost [7,9,10]. Therefore, the efficient control of the electrolyzer and compressor is essential to ensure economic feasibility [11].

Various HRS operation models that incorporate electrolyzers and compressors have been studied, with a focus on operating constraints and cost structures. A photovoltaic (PV)-based on-site HRS operation model was proposed in [12]. This model prioritized the use of solar power to minimize the dependence on external electricity, and power was drawn from a grid only when solar generation was insufficient. A rule-based operation model for an HRS with PV generation was presented in [13]. This model scheduled hydrogen production using a daily demand-forecasting algorithm for the HRS and controlled the compressor when the hydrogen storage tank level was below a certain threshold. An HRS configuration consisting of three storage tanks was used in [14]. Hydrogen production was initiated when the pressure in one of the tanks dropped below the maximum storage pressure, and the produced hydrogen was immediately compressed and stored. An optimization method for the electrolyzer operation in an HRS without renewable energy was proposed in [15]. The optimization method considering installation and scheduling with load (OPT-ISL) model minimized the total cost, including capital investment and operating expenses, with a focus on electrolyzer scheduling based on daily demand. In [16], the operation of an electrolyzer was optimized to minimize the operating cost of an HRS. Hydrogen demand was estimated, and a mixed integer linear programming (MILP)-based approach was employed to reduce costs by leveraging the peak–valley differences in electricity prices. In [17], hydrogen was externally supplied via trailers, and an optimization-based operation model was proposed to minimize the number of compressor operations. Through compressor control, the hydrogen compression costs were reduced by 20–30% compared with conventional models. In [18,19], the energy consumption was minimized by optimizing compressor operation according to hydrogen demand. The proposed models utilized a cascade hydrogen compression and storage system consisting of low-pressure, medium-pressure, and high-pressure stages. An optimized operation algorithm based on HRS demand forecasting was proposed in [20]. The algorithm estimated the next-day hydrogen demand distribution and scheduled hydrogen production based on the renewable energy availability.

Prior research has also emphasized optimizing electrolyzer operation within HRSs to reduce LCOH and operating costs [3–6]. However, the role of compressor control has not been examined in detail [15,16,18–20]. In most studies, compressor operation is based on predefined pressure thresholds [13,14,17]. However, to effectively reduce the overall operating cost of HRSs, an operation strategy that integrates electrolyzer and compressor control is required. Electrolyzer operation can lead to membrane damage and catalyst degradation, resulting in performance deterioration. In particular, frequent on/off switching during operation can accelerate degradation [21–23]. This degradation can reduce hydrogen production and increase electricity consumption, thus highlighting the requirement for an operation model that considers these factors.

This paper proposes an optimal operation model for a renewable-energy-coupled on-site HRS equipped with an energy storage system (ESS) that considers hydrogen production and compression costs. The proposed model aims to maximize profit by considering the hydrogen sales revenue, electricity purchase costs for any shortfall, and ESS degradation costs. The hydrogen production is estimated using a voltage equation. The power consumption of the compressor is modeled using the shaft power equation, which represents the mechanical power required for actual operation and includes ideal compression work and electromechanical losses [20]. Voltage degradation is calculated using an empirical model to analyze the impact of the electrolyzer operation and on/off cycling of the electrolyzer voltage. The proposed operation strategy is formulated as an MILP-based model, and its economic effectiveness is evaluated by comparing it with a conventional rule-based control method.

The remainder of this paper is organized as follows: Section 2 describes the configuration of the proposed on-site HRS and the mathematical modeling of each system component. Section 3 presents the simulation environment and assumptions. Section 4 describes the effectiveness of the proposed model through comparison with the rule-based control method using actual HRS operation data. Finally, Section 5 concludes the study.

2. On-Site HRS System

2.1. Architecture of the Proposed Operation Model

The proposed on-site HRS model consists of a PV system, ESS, proton exchange membrane water electrolyzer (PEMWE), compressor, and storage tanks. Figure 1 illustrates the architecture of the on-site HRS, where the black and blue lines represent electricity and hydrogen flow, respectively.

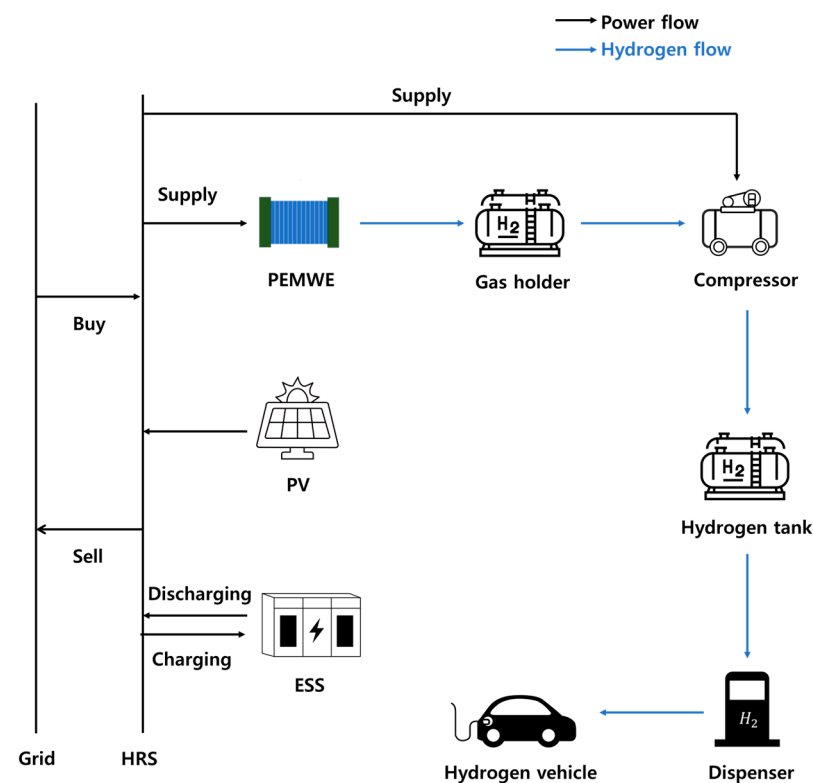


Figure 1. On-site HRS architecture.

The on-site HRS does not receive hydrogen from external sources but through on-site production. Hydrogen production and compression rely on PV power as the main electricity source. Surplus PV power is either stored in the ESS or sold to the grid. The stored ESS energy is used during nighttime or cloudy periods when PV generation is insufficient to ensure stable hydrogen production and compression. If PV and ESS power are insufficient, electricity can be supplied from the grid. The hydrogen compressor system is divided into low-pressure, medium-pressure, and high-pressure stages. Hydrogen is sequentially compressed through each stage via the corresponding storage tanks. The storage tanks are configured to low-pressure, medium-pressure, and high-pressure levels. A holder is placed between the electrolyzer and low-pressure compressor to ensure the stable storage and compression of the produced hydrogen.

2.2. Structure of the Proposed Operation Model

This section presents the structure of the HRS system model, including hydrogen production via electrolysis, storage-tank pressure calculation, and compressor modeling, as

illustrated in Figure 2. The hydrogen production model calculates the amount of hydrogen produced by the PEMWE based on the current and Faraday efficiency [24]. The amount of electricity required for hydrogen production is estimated using the voltage model [25]. The hydrogen tank capacity model estimates the pressure in the storage tanks using the hydrogen production calculated from the hydrogen production model and the output flow rate obtained from the compressor model. After the storage tank pressure is determined, the inlet and outlet pressures for each compressor stage are calculated and applied to the compressor model. The compressed hydrogen flows into the storage tanks. The compressor power consumption model calculates the power usage of the compressors based on the inlet/outlet pressures and other parameters. Finally, the degradation model evaluates the voltage degradation of the PEMWE system at each time step by considering the operating pattern and on/off cycling. The evaluated degradation value is fed back into the voltage model to enable the overall system model to account for the performance degradation of the electrolyzer.

The system modeling is described in Section 3.

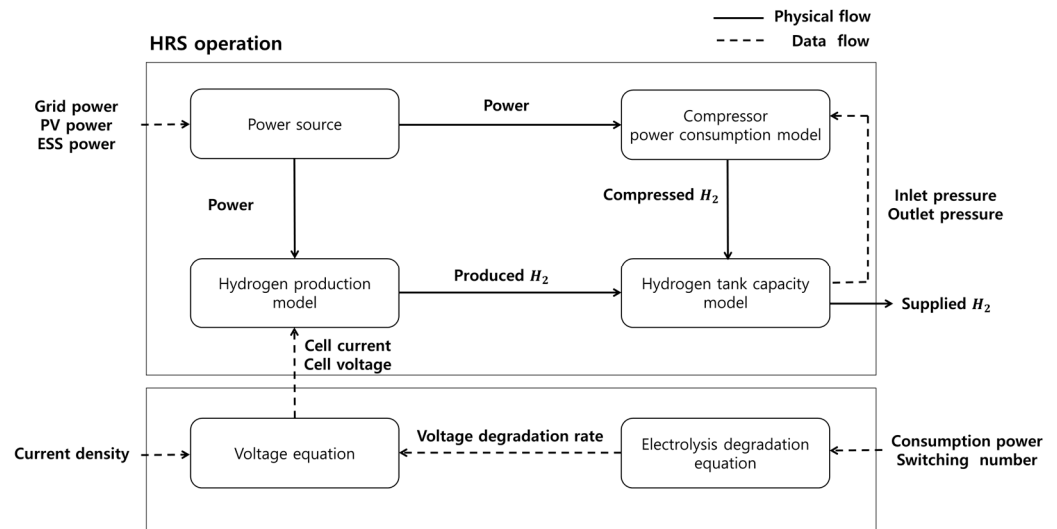


Figure 2. Physical and data flow structure of the proposed HRS operation model.

3. Hydrogen Production Model

3.1. Estimation of Amount of Hydrogen Produced

The amount of hydrogen produced can be estimated based on the electrolysis current and Faraday efficiency, as expressed in Equation (1) [26]. Electrolysis current I_t^{PEM} is calculated based on the cell current and active area, as shown in Equation (2). It is used with the voltage to compute the power consumption for hydrogen production, as shown in Equation (3). The electrolysis voltage (V_t^{PEM}) is determined by the cell voltage (V_t^{cell}) and number of cells (n^{cell}), as expressed in Equation (4). V_t^{cell} is influenced by operating conditions such as the current and temperature [27], and it directly affects the electrolysis power consumption (W_t^{PEM}). The method for estimating V_t^{cell} , which accounts for these influencing factors, is described in Section 3.2.

$$m_t^{H_2} = \eta^F \frac{I_t^{PEM}}{2F} \quad (1)$$

$$I_t^{PEM} = i_t^{cell} \cdot A \quad (2)$$

$$W_t^{PEM} = V_t^{PEM} \cdot I_t^{PEM} \quad (3)$$

$$V_t^{PEM} = V_t^{cell} \cdot n^{cell} \quad (4)$$

here, A represents the cell active area, $m_t^{H_2}$ denotes the amount of hydrogen produced, and η^F denotes the Faraday efficiency.

3.2. Estimation of Cell Voltage

The PEMWE voltage can be divided into the open-circuit voltage, activation, diffusion, and ohmic overpotentials, and it is expressed by Equation (5) [28].

$$V_t^{cell} = V^{ocv} + V_t^{diff} + V_t^{act} + V_t^{ohm} \quad (5)$$

The open-circuit voltage (V^{ocv}) represents the minimum voltage required to initiate the electrolysis reaction, and it is defined by the Nernst equation given by Equation (6) [28].

$$V^{ocv} = V^0 + \frac{RT}{2F} \left[\ln \left(\frac{p^{H_2} \sqrt{p^{O_2}}}{a^{H_2O}} \right) \right] \quad (6)$$

here, V^0 represents the reversible cell potential, which can be approximated as 1.23 V under standard temperature and pressure conditions [29]. R is the gas constant, T is the temperature, T^{ref} is the reference temperature, F is the Faraday constant, p^{H_2} and p^{O_2} are the partial pressures of hydrogen and oxygen, respectively, and a^{H_2O} denotes the water activity.

V_t^{diff} represents the overpotential caused by changes in the concentration of reactants on the electrode surface, and it is calculated using the Nernst equation given by Equation (7) [30].

$$V_t^{diff} = \frac{RT}{4F} \ln \left(\frac{\frac{P^a}{RT} \frac{n^{H_2}}{n^{H_2} + n^{H_2O,a}} + \frac{\delta^{e,a}}{D^{eff,a}} \frac{i_t}{4F}}{C_{O_2,m^0}} \right) + \frac{RT}{2F} \ln \left(\frac{\frac{P^c}{RT} \frac{n^{O_2}}{n^{O_2} + n^{H_2O,c}} + \frac{\delta^{e,c}}{D^{eff,c}} \frac{i_t}{4F}}{C_{H_2,m^0}} \right) \quad (7)$$

here, j_t is current density, $C_t^{O_2,m}$ and $C_t^{H_2,m}$ are the oxygen and hydrogen concentration at the membrane electrode interface, respectively, C_{O_2,m^0} and C_{H_2,m^0} are the reference working conditions, $\delta^{e,a}$ and $\delta^{e,c}$ are the anode and cathode thicknesses, respectively, $D^{eff,a}$ and $D^{eff,c}$ denote the effective binary diffusivities, and P^a and P^c denote the operating pressures at the anode and cathode, respectively. $\dot{n}_t^{O_2}$, $\dot{n}_t^{H_2}$, and $\dot{n}_t^{H_2O}$ denote the molar flows of oxygen, hydrogen, and water, respectively.

The losses that occur during electrochemical activation are represented by V^{act} and are based on the Butler–Volmer equation. V_t^{act} is expressed as the sum of the activation overvoltages at the anode and cathode and the voltage degradation caused by on/off switching, as shown in Equation (8).

$$V_t^{act} = \left(\frac{RT}{\alpha^a F} \ln \left[\frac{i_t}{2i_{0,a}} + \sqrt{1 + \left(\frac{i_t}{2i_{0,a}} \right)^2} \right] \right) + V_t^{act,on} \quad (8)$$

here, α^a and α^c are the charge transfer coefficients of the anode and cathode, respectively, $j_{0,a}$ and $j_{0,c}$ represent the exchange current densities of the anode and cathode, respectively.

The ohmic overvoltage varies depending on the electrical resistance caused by the transport of electrons and ions, and it is based on Ohm's law. V_t^{ohm} is expressed as a function of the membrane electrode assembly (MEA) resistance and current density, as given by Equation (9).

$$V_t^{ohm} = i_t \cdot \left(\frac{\delta_t^{mem}}{\sigma_t^{mem}} \right) \quad (9)$$

here, δ_t^{mem} represents the membrane thickness and σ_0^{mem} denotes the initial conductivity.

3.3. Hydrogen Compressor Model

The compressor is modeled assuming that multiple pistons operate in phase to maintain a constant mass flow rate when the compressor is on, and its performance is affected by the inlet pressure [31]. The hydrogen compressor is modeled with reasonable accuracy by assuming that hydrogen is an ideal gas, and the process is considered to be a polytropic compression [32]. The maximum amount of hydrogen that each compressor can compress at time t is given by Equation (10), and this amount depends on the inlet pressure. The power consumption of the compressor is modeled based on the shaft power equation given by Equation (11). $W_{t,i}^{comp}$ represents the power consumption of the compressor, which is affected by the inlet and outlet pressures.

$$m_{t,j}^{comp} = N_j \times V_j^{disp} \times \eta_j^{comp} \times \frac{P_{t,j}^{inlet}}{zRT} \quad (10)$$

$$W_{t,j}^{comp} = C^p \times \frac{T}{\eta_j^{comp}} \left(\left(\frac{P_{t,j}^{outlet}}{P_{t,j}^{inlet}} \right)^{\frac{r-1}{r}} - 1 \right) \times m_{t,j}^{comp} \quad (11)$$

here, j represents the number of compressors, $m_{t,i}^{comp}$ is the volumetric gas flow rate of the compressor, N_i is the compressor speed, V_i^{disp} is the displacement of the compressor, $\eta_{t,i}^{comp}$ is the volumetric efficiency, $P_{t,i}^{outlet}$ and $P_{t,i}^{inlet}$ are the outlet and inlet pressures, respectively, C^p represents the specific heat of hydrogen, z is the compressibility factor of hydrogen, and r is the heat capacity ratio, which is assumed to be constant.

3.4. HRS Operation

The objective of the proposed on-site HRS is to maximize the operating profit, and the objective function is given by Equation (12). The operating profit consists of the electricity sales revenue, which refers to the income generated by feeding surplus power back to the grid, and the hydrogen sales revenue, which refers to the income earned by supplying hydrogen to meet a specified demand. The operating costs include the electricity purchase cost, which is incurred to procure the power required in addition to the power supplied by renewable energy, and the ESS degradation cost (Equation (13)), which accounts for the wear and tear associated with charging and discharging operations.

Equation (14) expresses the power balance in the HRS. In addition, Equations (15) and (16) define the power-exchange constraints that prevent simultaneous electricity purchase and sale at the same time step.

$$\text{Maximize} \sum_{t=7}^N \left(W_t^{grid,sell} \cdot c^{smp} + H_t^{sell} \cdot c^{H_2} \right) - \left(W_t^{grid,buy} \cdot c^{grid} + C_t^{ess_wear} \right) \quad (12)$$

$$C_t^{ess_wear} = c^{wear} \left(W_t^{ESS,ED} + W_t^{ESS,EC} \right) \quad (13)$$

$$W_t^{PEM} + W_t^{comp} + W_t^{grid,sell} + W_t^{ESS,EC} = W_t^{pv} + W_t^{ESS,ED} + W_t^{grid,buy} \quad (14)$$

$$W_t^{grid,buy} = W_t^{grid,buy,max} \times U_t^{resold} \quad (15)$$

$$W_t^{grid,sell} = W_t^{grid,sell,max} \times (1 - U_t^{resold}) \quad (16)$$

here, $W_t^{grid,sell}$ is the exported electricity, c^{smp} is the unit price of the electricity sold to the grid, H_t^{sell} is the amount of hydrogen sold, and c^{H_2} is the unit price of hydrogen. $W_t^{grid,buy}$ is the electricity purchased from the grid, and c^{grid} is its unit price. $C_t^{ess_wear}$ represents the ESS degradation cost. t indicates the time step of the HRS operation, and N denotes the set of

operating periods. c^{wear} is the unit degradation cost of the ESS, and $W_t^{ESS,ED}$ and $W_t^{ESS,EC}$ represent the discharging and charging powers of the ESS, respectively. W_t^{PEM} is the electricity used for hydrogen production via electrolysis, W_t^{comp} is the power consumption of the compressor, and W_t^{pv} is the generated PV power.

To account for ESS degradation costs, the average wear cost (AWC) was estimated based on the depth of discharge (DoD)–achievable cycle count (ACC) relationship, as expressed in Equation (17) [33]. This linear approximation—derived by linearizing the battery degradation model—was adopted to ensure computational efficiency, preserve the consistent linearity of the proposed optimization framework, and enhance practical implementability. The ESS degradation cost is calculated based on a fixed DoD and ACC corresponding to a specific end-of-life (EoL) criterion. DoD represents the depth to which the battery is discharged, and ACC refers to the number of charge/discharge cycles that can be repeated at that DoD. Battery aging depends on the DoD. To quantify this effect, this study applies a lifetime evaluation model that reflects the DoD–ACC relationship of lithium-ion batteries. The battery degradation cost (c^{wear}) is calculated by dividing the battery price by the total usable energy, where *Efficiency* accounts for the losses occurring during the charging and discharging processes.

$$c^{wear} = \frac{\text{Battery Price}}{\text{ACC}(D) \times 2 \times \text{DoD} \times \text{Battery size} \times \text{Efficiency}} \quad (17)$$

To prevent the simultaneous charging and discharging of the hydrogen storage tanks, adjacent compressors are not allowed to operate simultaneously. The compressor operation is modeled using binary variables, as given by Equations (18) and (19).

$$m_{t,i}^{comp} \leq U_{t,i}^{comp} \times m_{t,i}^{comp,max} \quad (18)$$

$$U_{t,i}^{comp} + U_{t,i+1}^{comp} \leq 1 \quad (19)$$

here, $U_{t,i}^{comp}$ is a binary variable that represents the operating status of the compressor and $m_{t,i}^{comp,max}$ denotes the maximum compression capacity of the compressor.

The hydrogen storage tanks were modeled using the ideal gas law. Within the considered pressure range, the effects of temperature and the real-gas compressibility factor (z) were assumed negligible and were therefore excluded from the model [34]. The amount of hydrogen produced via electrolysis is assumed to be equal to the hydrogen that flows into the holder of tank 1. The low-pressure, medium-pressure, and high-pressure tanks receive the hydrogen that has been compressed from the lower-pressure stage, as expressed in Equation (20). The outflow from the hydrogen tanks is categorized as either the inflow to the next-stage compressor or the final outflow for hydrogen sales, as given by Equation (21). Equation (22) represents the hydrogen storage state of the tanks, and Equation (23) defines the initial and final constraints of the hydrogen storage tanks.

$$m_{t,b}^{tank,in} = \begin{cases} m_t^{H_2} & \text{if } b = 1 \\ m_{t,j}^{comp} & \text{if } b > 1 \end{cases} \quad (20)$$

$$m_{t,b}^{tank,out} = \begin{cases} m_{t,j+1}^{comp} & \text{if } b = 1, 2, 3 \\ H_t^{sell} & \text{if } b = 4 \end{cases} \quad (21)$$

$$P_{t,b}^{tank} = \frac{RT}{MV3600} \left(m_{t-1,b}^{tank} + \left(m_{t,b}^{tank,in} - m_{t,b}^{tank,out} \right) \right) \frac{z}{1000} \quad (22)$$

$$P_{1,b}^{tank} = P_{t^{fin},b}^{tank} \quad (23)$$

The hydrogen storage tanks are represented in terms of pressure, where b denotes the tank index. $m_t^{tank,in}$ and $m_t^{tank,out}$ represent the hydrogen inflow and outflow of a tank, respectively. M is the molar mass of hydrogen, and V is the volume of the tank [35].

The operating and state-of-charge (SOC) constraints for the ESS operation are defined by Equations (24)–(29).

$$SOC_1 = SOC_{t_{fin}} \quad (24)$$

$$SOC^{min} \leq SOC_t \leq SOC^{max} \quad (25)$$

$$SOC_t = SOC_{t-1} + (W^{ESS,EC} \cdot \eta^{ESS,EC} - W^{ESS,ED} / \eta^{ESS,ED}) \quad (26)$$

$$W^{ESS,EC} \leq U_t^{ESS,EC} W^{ESS,EC,max} \quad (27)$$

$$W^{ESS,ED} \leq U_t^{ESS,ED} W^{ESS,ED,max} \quad (28)$$

$$U_t^{ESS,EC} + U_t^{ESS,ED} \leq 1 \quad (29)$$

The initial and final SOC of the ESS are set to be equal through Equation (24). The upper and lower bounds of the battery SOC are constrained within SOC^{min} and SOC^{max} , respectively, as defined by Equation (25). The SOC at time t is determined by the amount of charging and discharging, as expressed in Equation (26). Equations (27) and (28) limit the charging and discharging powers of the ESS, whereas Equation (29) imposes a constraint to prevent simultaneous charging and discharging.

3.5. PEMWE Degradation Calculation

The degradation of the PEMWE considers the membrane damage and catalyst degradation that occur during operation. On/off switching significantly affects catalyst degradation, leading to an increase in V^{act} . An empirical model derived from the relationship between voltage and internal resistance is applied to quantitatively assess the effect on the catalyst. In this model, the level of degradation is calculated based on the number of on/off cycles and curve-fitting coefficient k^{on} . Equation (30) represents the degradation caused by on/off switching [36].

$$V_c^{act,on} = k^{on} \ln(c) + V_{t-1}^{act,on} \quad (30)$$

MEA degradation is categorized into membrane thickness degradation and conductivity degradation. The reduction in the membrane thickness is affected by the hydroxyl radicals generated during hydrogen production, and the formation of these radicals is determined by the system temperature and current density. Membrane thickness degradation is estimated based on the amount of fluorine released. The conductivity of the MEA is determined by its thickness. Equations (31) and (32) represent the degradation in the membrane thickness and MEA conductivity, respectively [37].

$$\delta_t = \delta_0 - \frac{\sum_{t=7}^N FDR_t}{\lambda^{mem} \rho^{mem}} \quad (31)$$

$$\sigma_t = \sigma_0 \left(\frac{\delta_t}{\delta_0} \right)^2 \quad (32)$$

Here, δ represents the MEA thickness, λ^{mem} is the fluorine content of the MEA, ρ^{mem} is the MEA density, FDR denotes the fluorine discharge rate, and σ represents the MEA conductivity.

4. Case Studies and Discussion

4.1. Case Study Setting

Comparative analysis is conducted for three cases to evaluate the effectiveness of the proposed model, as shown in Table 1. Case 1 controls both the compressor and electrolyzer through a predefined simple control. In this context, the simple control method involves controlling the compressor and electrolyzer based on the pressure levels of the hydrogen storage tanks. When the pressure in the storage tanks falls below a certain threshold, the compressor is activated to maintain the hydrogen level above a predefined minimum [20].

Table 1. Operation methods for different cases.

Case	Electrolysis Control	Compressor Control
Case 1	simple control	simple control
Case 2	optimal control	simple control
Case 3	optimal control	optimal control

In Case 2, the compressor operates under a simple control strategy, while the electrolyzer is included in the optimal control framework. In Case 3, both the compressor and the electrolyzer are incorporated into the optimization model. The detailed relationship between the inlet and outlet pressures and the power consumption of the compressor is provided in Appendix A.

The simulation is based on real-world HRS demand and PV generation data. The HRS demand data were collected from the Hydrogen Distribution Information System [38], while the PV generation data were obtained from a specific region in Republic of Korea. The optimization model is solved using the Gurobi 12.0.3 solver, the simulation is conducted using a 15 min time step over a 4-week period. The target HRS operates from Monday to Saturday, as there is no demand for hydrogen on Sundays; hence, daily and weekend patterns are incorporated in the model. It is assumed that the storage tanks maintain a minimum hydrogen level at the end of Saturday and start at the same level on Monday. Table 2 provides the simulation parameters, such as grid electricity purchase and feed-in tariffs and the capacities of the PV system, ESS, and PEMWE. The feed-in electricity price is assumed to be 50% of the system marginal price. The electricity and hydrogen prices used in the simulation were set according to the Korean market environment.

Table 2. Simulation parameters.

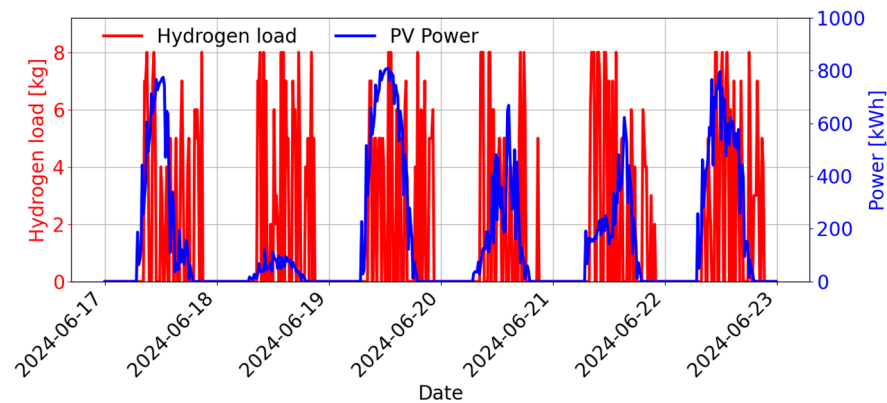
Name	Value	Name	Value
PV	1 MW	feed-in tariff	56.7 KRW/kWh
PEMWE	500 kW	electricity purchase price	TOU price
ESS	1 MW	hydrogen selling price	8800 KRW/kg
SOC range	20–80%	ESS capacity cost	110,694 KRW/kW

The initial and constant parameters used for the PEMWE modeling are based on previous studies [39–41]. In the case study, the MEA is assumed to be Nafion 117, and the system current density range is set as 0–2 A/cm². The maximum number of PEMWE cycles is assumed to be 5000. The maximum hydrogen storage capacity of the storage tanks is defined by the volume and maximum pressure, as shown in Table 3. The gas holder was assumed to have sufficient capacity to store the hydrogen produced by the PEMWE, and its maximum pressure was set at 10 bar.

Table 3. Specifications of hydrogen storage tanks.

Tank	Volume (m ³)	$p_{\text{tank,max}}$ (bar)
Low-pressure tank	111.35	40
Medium-pressure tank	7.42	550
High-pressure tank	3.09	960

The PV generation and HRS demand data used in the case study are based on 15 min interval input data. The hydrogen demand is derived from pressure variations in the trailer, which are converted into mass using the ideal gas law. Figure 3 shows the hydrogen demand and PV generation data used in the case study.

**Figure 3.** HRS hydrogen demand and PV generation.

4.2. Results and Discussion

The simulation results (Table 4) demonstrate the differences in the revenue and cost structures between the operation models. Although all the cases generate the same hydrogen sales revenue, there are noticeable differences in terms of the electricity sales revenue, electricity purchase cost, and ESS degradation cost.

Table 4. Operating profit for different cases.

Category		Case 1	Case 2	Case 3
Revenue	Hydrogen sales revenue [KRW]	28,107,200	28,107,200	28,107,200
	Electricity sales revenue [KRW]	14,671,118	10,455,947	10,169,874
	Electricity purchase cost [KRW]	17,007,768	2,452,242	1,332,761
Cost	Basic service charge [KRW]	691,386	96,420	89,900
	ESS wear cost [KRW]	1,600,636	156,484	113,457
Profit	Total Profit [KRW]	23,478,528	35,858,001	36,740,957

Case 1 shows the highest electricity sales revenue. However, this is offset by the highest electricity purchase cost, resulting in a limited increase in the net profit. Case 2 achieves a higher operating profit compared with Case 1. However, the electricity purchase cost for Case 2 is more than 80% higher than that for Case 3, leading to a relatively lower total profit. In contrast, Case 3 shows the lowest electricity sales revenue but achieves the highest reductions in the electricity purchase cost and ESS degradation cost, resulting in the highest overall cost efficiency. These results indicate that an optimal HRS operation model that comprehensively considers components, such as the electrolyzer and compressor, effectively improves the economy and profitability of HRS operations.

Table 5 presents a comparison of the power consumption characteristics of the electrolyzer and compressor systems and the renewable energy self-consumption rate. Case 1

exhibits the highest power consumption and lowest self-consumption rate of renewable energy. Case 2 shows less power consumption and more self-consumption compared with Case 1 but more power usage and less self-consumption compared with Case 3. Case 3 manages the power consumption in the most effective manner through the optimized operation of the compressor and electrolyzer systems. Its renewable energy self-consumption rate is approximately 40.1% higher than that for Case 1 and 2.4% higher than that for Case 2. These results demonstrate that an optimal HRS operation model that integrates the electrolyzer and compressor operations effectively increases renewable energy utilization and decreases electricity consumption.

Table 5. Power consumption for different cases.

Category	Case 1	Case 2	Case 3
Hydrogen production and compression [kg]	3194	3194	3194
Power consumption for electrolysis [kWh]	236,860	234,570	234,459
Power consumption for compression [kWh]	6606	6638	6521
PV self-consumption [kWh]	134,330	223,850	229,351
PV power export to grid [kWh]	258,750	184,407	179,363
ESS discharge [kWh]	15,186	1483	1076
Grid power consumption [kWh]	93,951	15,872	10,552
Renewable energy self-consumption rate [%]	55.14	92.86	95.22
Category	Case 1	Case 2	Case 3

Figure 4 shows the trends in the electricity sales (red lines) and purchases (blue lines) for each case. Case 1 shows frequent electricity purchases and the highest total electricity purchased during the simulation period. Case 2 shows a reduced frequency of electricity sales and purchases compared with Case 1. The electricity purchases are higher compared with Case 3 because of the limitations in efficient renewable energy utilization due to the simple control of the compressor, which results in higher dependency on grid electricity. Case 3 shows the lowest electricity sales and purchases, thereby demonstrating that the renewable-energy-based optimal operation model effectively reduces electricity costs.

Figures 5 and 6 show the trends of the power consumption of the compressor and capacity of the hydrogen storage tanks, respectively. In Cases 1 and 2, the storage tank levels remain relatively constant owing to the rule-based control. In particular, the low-pressure and medium-pressure tanks maintain stable levels with minimal fluctuations, and the other tanks follow similar patterns. This indicates a limitation of the rule-based operation, where the compressor is activated based on predefined tank capacity thresholds. As a result, during periods of high variability in hydrogen demand, the frequency of compressor operation increases, leading to higher power consumption. This is also evident in Figure 5. In contrast, Case 3 employs an optimization-based operation strategy that allows the hydrogen storage levels to be flexibly adjusted in response to demand changes. Consequently, the storage tanks maintain higher levels compared with the other cases. This reduces the frequency of compressor operation and the pressure ratio between the inlet and outlet, thereby reducing the power consumption of the compressor. This characteristic is prominent in most tanks except the holder tank. These results indicate that the optimized use of storage tanks and reduction in the frequency of compressor operation decreases the overall power consumption.

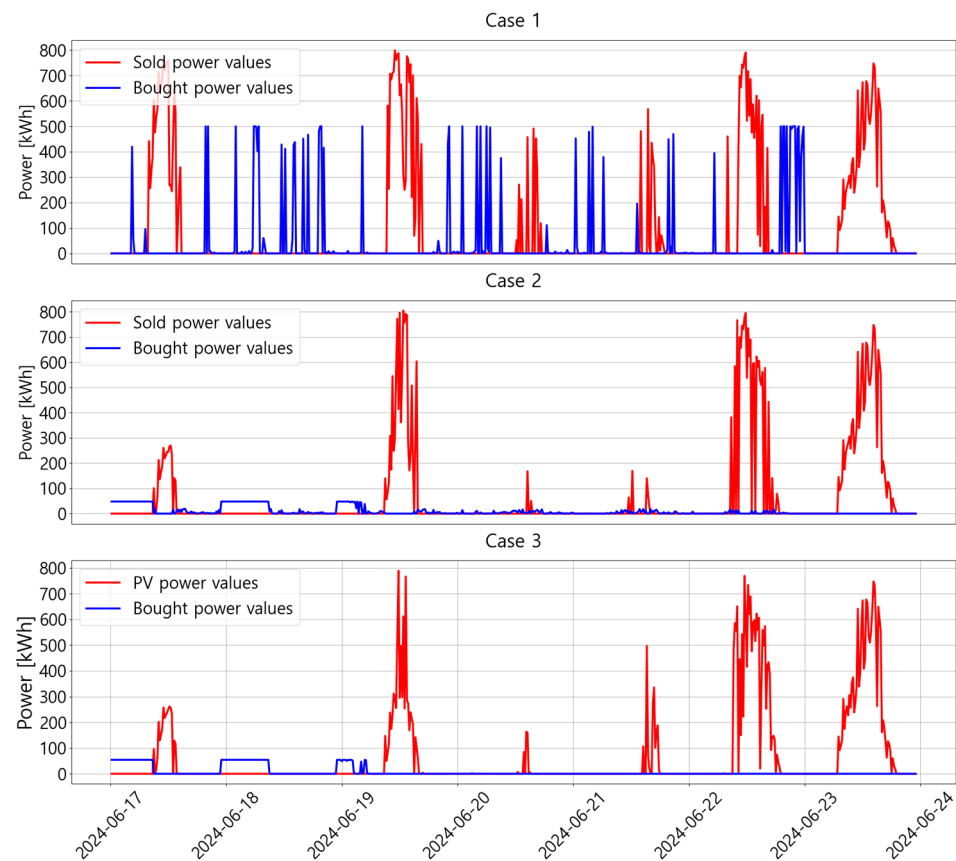


Figure 4. Electricity sale and purchase trends for different cases.

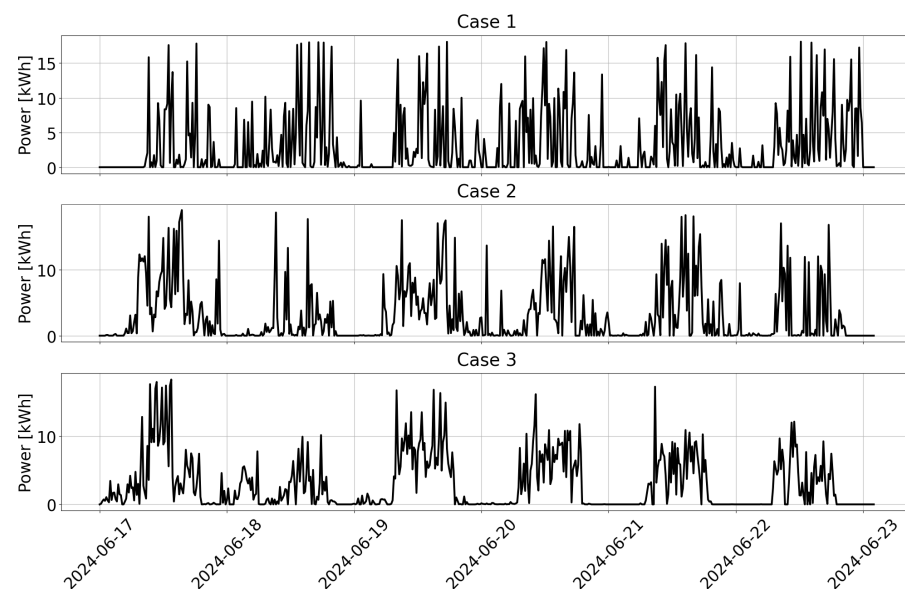


Figure 5. Power consumption of compressor for different cases.

A post-evaluation of the degradation in the electrolyzer performance is conducted based on the operational results. Case 1 exhibits the highest degradation, with a membrane thickness reduction of approximately 0.98% and 248 on/off cycles over one month. Based on this, the estimated MEA lifetime is approximately 4 years and 3 months. In contrast, Case 2 shows a membrane thickness reduction of approximately 0.87% with 100 on/off cycles, resulting in an estimated MEA lifetime of approximately 4 years and 8 months. Case 3 exhibits the lowest degradation, with a membrane thickness reduction of 0.84% and 93 on/off cycles, leading to an estimated MEA lifetime of approximately 4 years and

11 months. The lower reduction in the membrane thickness and improved MEA lifetime in Case 3 suggest that the differences in electrolyzer operation patterns based on the control model suppress the degradation in the electrolyzer performance. However, it should be noted that these results are based on short-term data obtained for approximately one month of operation, and the absolute differences in lifetime may be limited. Nevertheless, even small differences in degradation rates can accumulate during long-term operation. This can deteriorate the performance and reduce the lifespan of the electrolyzer, thereby highlighting the importance of the operation model. Although the simulation targets a single on-site HRS, the proposed framework can be applied to stations of varying sizes and load characteristics through appropriate parameter tuning.

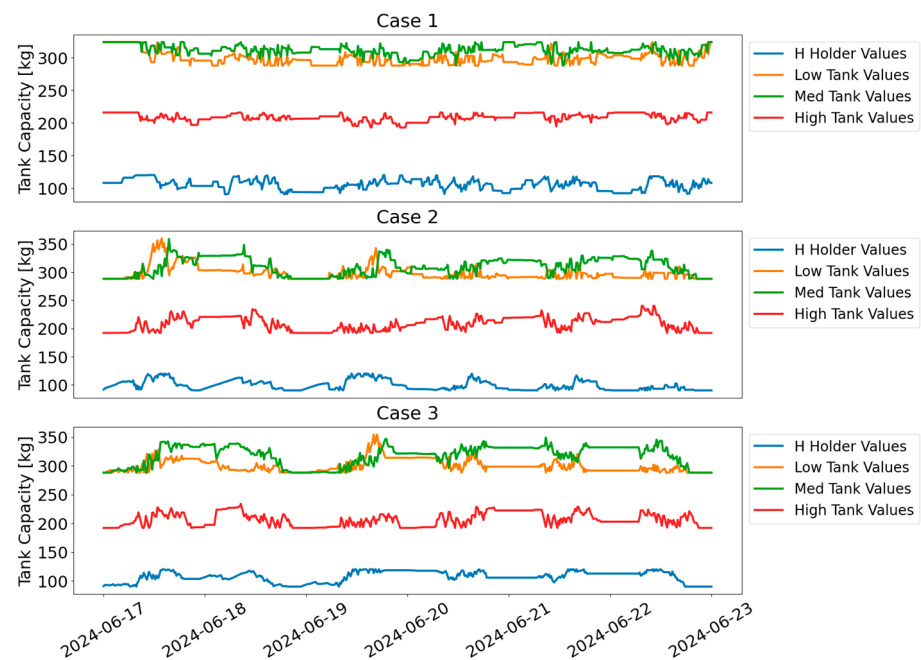


Figure 6. Capacity of hydrogen storage tanks for different cases.

5. Conclusions

This paper proposes an integrated simulation framework for an HRS that incorporates a lifetime evaluation model for electrolyzer systems. An optimal operation model is presented to maximize the operating profit in an on-site HRS by scheduling hydrogen production and compressor operation.

Simulation results show that Case 3, which applies the proposed model, achieves the highest reductions in the electricity purchase and ESS degradation costs, resulting in an increase of 56.49% in the total profit compared with Case 1 and 2.5% compared with Case 2. Additionally, by optimizing the power usage of the compressor and electrolyzer systems, the energy consumption is reduced by up to 1.02% and the renewable energy self-consumption rate is improved by up to 40.1%, thereby confirming the effectiveness of the model in enhancing energy efficiency. The electrolyzer also exhibits the lowest performance degradation, indicating that the proposed operation strategy is effective for extending equipment lifetime during long-term operation.

This study specifically evaluates the economic impacts of electrolyzer and compressor control and does not account for the effects of other auxiliary subsystems. To simplify the model, we fixed the temperatures of the compressor and storage tanks, applied ideal-gas and isentropic relationships, and represented the voltage rise due to electrolyzer on/off cycling in a simplified manner. These assumptions limit our ability to fully capture real-world efficiency losses and long-term degradation effects.

For future work, we will incorporate temperature-dependent characteristics to improve the accuracy of storage-tank mass estimation and compressor efficiency/power consumption prediction and explicitly integrate degradation effects into a long-horizon operational model to quantify cumulative impacts. We also plan to enhance the model's applicability and practical relevance through comparative analyses that include demand variability and cross-country tariff and policy frameworks.

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Abbreviations

The following abbreviations are used in this manuscript:

ACC	Achievable cycle count
AWC	Average wear cost
DoD	Depth of discharge
EoL	End-of-life
ESS	Energy storage system
FCEVs	Fuel-cell electric vehicles
HRS	Hydrogen refueling station
LCOH	Levelized cost of hydrogen
MEA	Membrane electrode assembly
MILP	Mixed integer linear programming
PEMWE	Proton exchange membrane water electrolyzer
PV	Photovoltaic
SOC	State-of-charge
A	Cell active area [m^2]
α^{H_2O}	Water activity
α^a	Anode charge transfer coefficient
α^c	Cathode charge transfer coefficient
C^{ess_wear}	ESS degradation cost [KRW/kWh]
c^{grid}	Unit price of grid electricity [KRW/kWh]
c^{H_2}	Unit price of hydrogen [KRW/kg]
c^{exp}	Unit price of exported electricity [KRW/kWh]
c^{wear}	Unit degradation cost [KRW/kWh]
$C_t^{O_2,m}$	Oxygen concentration [$\text{mol}\cdot\text{m}^{-3}$]
$C_t^{H_2,m}$	Hydrogen concentration [$\text{mol}\cdot\text{m}^{-3}$]

$D^{eff,a}$	Anode effective binary diffusivity [$\text{m}\cdot\text{s}^{-1}$]
$D^{eff,c}$	Cathode effective binary diffusivity [$\text{m}\cdot\text{s}^{-1}$]
$\delta^{e,a}$	Anode thickness [m]
$\delta^{e,c}$	Cathode thickness [m]
F	Faraday constant [$\text{C}\cdot\text{mol}^{-1}$]
C^p	Specific heat of hydrogen [$\text{J}\cdot\text{kg}^{-1}\cdot\text{K}^{-1}$]
I_t^{PEM}	Electrolysis current [A]
$j^{0,a}$	Anode exchange current density [$\text{A}\cdot\text{m}^{-2}$]
$j^{0,c}$	Cathode exchange current density [$\text{A}\cdot\text{m}^{-2}$]
j_t	Current density [$\text{A}\cdot\text{m}^{-2}$]
r	Heat capacity ratio
k^{on}	Number of on/off cycles and curve-fitting coefficient
λ^{mem}	Membrane water content
$m_t^{H_2}$	Amount of hydrogen produced [kg]
$m_t^{tank,in}$	Hydrogen inflow [$\text{kg}\cdot\text{s}^{-1}$]
$m_t^{tank,out}$	Hydrogen outflow [$\text{kg}\cdot\text{s}^{-1}$]
N_i	Compressor speed [RPM]
n^{cell}	Number of cells
$\eta_{t,i}^{comp}$	volumetric efficiency
p^a	Anode pressure [Pa]
p^c	Cathode pressure [Pa]
p_t^{outlet}	Outlet pressure [Pa]
$p_{t,i}^{inlet}$	Inlet pressure [Pa]
R	Gas constant [$\text{J}\cdot\text{mol}^{-1}\cdot\text{K}^{-1}$]
ρ^{mem}	Membrane density [$\text{kg}\cdot\text{m}^{-3}$]
σ_0^{mem}	Membrane conductivity [$\text{S}\cdot\text{m}^{-1}$]
SOC_t	ESS SOC
SOC^{min}	Minimum SOC constraints
SOC^{max}	Maximum SOC constraints
T	Temperature [K]
T^{ref}	Reference temperature [K]
$U_{t,i}^{comp}$	Compressor on/off status (binary)
V_t^{act}	Activation overvoltage [V]
V_t^{cell}	Cell voltage [V]
V_t^{diff}	Diffusion overpotential [V]
V_t^{ohm}	Ohmic voltage [V]
V_{ocv}	Open-circuit voltage [V]
V_t^{PEM}	Electrolysis voltage [V]
V_i^{disp}	Compressor displacement [$\text{m}^3\cdot\text{rev}^{-1}$]
W_t^{comp}	Compressor power [KWh]
$W_t^{ESS,EC}$	ESS charging power [KWh]
$W_t^{ESS,ED}$	ESS discharging power [KWh]
$W_t^{grid_buy}$	Electricity purchased from the grid [$\text{KRW}\cdot\text{KWh}^{-1}$]
$W_t^{grid_sell}$	Exported electricity [$\text{KRW}\cdot\text{KWh}^{-1}$]
W_t^{PEM}	electrolysis power consumption [KWh]
W_t^{pv}	PV power [KWh]
z	Compressibility factor of hydrogen
$\dot{n}_t^{H_2}$	Molar flows of hydrogen [$\text{mol}\cdot\text{s}^{-1}$]
$\dot{n}_t^{H_2O}$	Molar flows of water [$\text{mol}\cdot\text{s}^{-1}$]
$\dot{n}_t^{O_2}$	Molar flows of oxygen [$\text{mol}\cdot\text{s}^{-1}$]
λ^{mem}	Fluorine content of the MEA

Appendix A

The power consumption of each compressor is derived using a lookup table based on the inlet and outlet pressures, as listed in Tables A1–A3. In each table, the vertical and horizontal axes represent the compressor inlet and outlet pressures, respectively. The pressures are expressed in terms of a pressure ratio. The power consumption of the compressor increases with the pressure ratio.

Table A1. Lookup table for low-pressure compressor [kWh].

Pressure [bar]	30	31	32	...	38	39	40
1	0.0881	0.0893	0.0906	...	0.0979	0.0991	0.1002
2	0.0603	0.0614	0.0626	...	0.0707	0.0716	0.0725
3	0.0478	0.0489	0.0518	...	0.0571	0.0579	0.0588
⋮	⋮	⋮	⋮	...	⋮	⋮	⋮
8	0.0246	0.0253	0.0261	...	0.0301	0.0307	0.0313
9	0.0227	0.0234	0.0241	...	0.0273	0.0279	0.0285
10	0.0197	0.0204	0.0211	...	0.0249	0.0254	0.0261

Table A2. Lookup table for medium-pressure compressor [kWh].

Pressure [bar]	500	510	520	...	580	590	600
30	0.0660	0.0668	0.0674	...	0.0712	0.0718	0.0724
31	0.0649	0.0656	0.0663	...	0.0701	0.0707	0.0713
32	0.0639	0.0645	0.0652	...	0.0690	0.0696	0.0702
⋮	⋮	⋮	⋮	...	⋮	⋮	⋮
38	0.0582	0.0589	0.0595	...	0.0631	0.0637	0.0643
39	0.0574	0.0581	0.0587	...	0.0622	0.0628	0.0635
40	0.0567	0.0573	0.0580	...	0.0615	0.0621	0.0627

Table A3. Lookup table for high-pressure compressor [kWh].

Pressure [bar]	750	760	770	...	930	940	950
500	0.00658	0.00681	0.00726	...	0.0106	0.00108	0.012
510	0.00624	0.00647	0.00692	...	0.01024	0.01043	0.01062
520	0.00591	0.00614	0.00659	...	0.00989	0.01008	0.01027
⋮	⋮	⋮	⋮	...	⋮	⋮	⋮
580	0.00408	0.00431	0.00473	...	0.00794	0.00812	0.0083
590	0.00380	0.00402	0.00445	...	0.00764	0.00782	0.0080
600	0.00353	0.00375	0.00417	...	0.00734	0.00753	0.0077

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