4E Analysis of LNG Cold Energy Integration for Sustainable Green Ammonia Production

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**Abstract.** The increase in average global temperatures emphasizes the urgency of adopting new and renewable energy sources to mitigate carbon emissions. Among these, ammonia emerges as a promising clean and renewable energy carrier for the future. This study proposes a green ammonia production scheme based on alkaline water electrolysis (AWE) powered by solar photovoltaics and integrated with LNG terminals to harness cold energy for the cryogenic air separation unit and ammonia liquefaction. The proposed scheme is simulated using Aspen Plus software to determine mass and energy balances, with subsequent analyses evaluating energy, exergy, environmental, and economic (4E) aspects. The results indicate an energy efficiency of 55.24% and an exergy efficiency of 96.54%. The exergy analysis identifies a total exergy destruction rate of 678,355 kW, with the Ammonia Synthesis Unit contributing 79.54%, highlighting the need to improve energy efficiency in this process. The integration of LNG cold energy enhances the energy efficiency of the system by approximately 3%. Environmentally, the proposed scheme achieves zero direct onsite carbon emissions, affirming its green credentials. Economically, the levelized cost of ammonia (LCOA) is calculated to be 469 $/tNH3, which reduces to 250 $/tNH3 when accounting for the sale of oxygen as a by-product. These findings underscore the viability and competitiveness of green ammonia production compared to conventional ammonia production methods based on steam methane reforming.

# INTRODUCTION

The emphasis on greenhouse gas (GHG) emissions reduction is a key priority across multiple sectors, with the energy sector being the largest contributor, followed by agriculture and industrial processes [1]. As part of the efforts to curb emissions within the energy sector, projections indicate that Electric Power and hydrogen will experience the highest compounded annual growth rates, with hydrogen growing at 6.5% annually until 2050, reflecting its growing importance as a clean energy source [2]. Hydrogen can be synthesized from various raw materials and energy sources, and it holds significant promise as an energy carrier. Green hydrogen, in particular, offers a sustainable and zero-emission solution, as it is produced from renewable energy sources and can be utilized across a wide range of sectors [3].

In terms of practical applications, hydrogen can be used directly as a fuel for vehicles, an industrial raw material, and a fuel for power generation. While direct electrification powered by renewable energy is considered the most efficient and cost-effective solution for transportation, hydrogen offers a flexible alternative. It can be stored for later use, addressing the challenge of energy storage. However, hydrogen transportation presents logistical difficulties, making it easier to transport in forms such as methanol, ammonia, or dibenzyl toluene. Ammonia, in particular, has emerged as a promising carrier for hydrogen storage and transportation due to its ease of transport and storage when produced from renewable energy sources [5]. Furthermore, ammonia is being explored as a co-firing material in existing power plants. In Indonesia, while co-firing ammonia in combined-cycle gas turbine (CCGT) power plants does not lead to significant reductions in CO2 emissions, its application in coal-fired power plants (CFPP) could produce substantial emission reductions due to the larger share of CFPP in the country’s energy mix [7].

Ammonia is also being investigated for use as an energy source in fuel cells, transportation, industrial processes, and power generation [8]. One of the key advantages of ammonia as an energy carrier is its lower transportation cost compared to LNG, approximately 47% cheaper, with respective costs of $0.02/kg for ammonia versus $0.038/kg for LNG. However, ammonia’s lower energy density does result in transportation costs being around 45% higher than LNG [9]. Despite this, ammonia is more manageable than hydrogen because it can exist as a liquid at -33°C at atmospheric pressure or at room temperature under moderate pressure (around 10 bar). Liquid ammonia contains a high percentage of hydrogen by weight (17.8%) and has an energy density of 13.77 MJ/L at 20°C and 8.6 bar, making it more stable and easier to transport than hydrogen or methanol [10]. Its energy density is about half that of gasoline and ten times greater than that of batteries, and it also has lower storage costs compared to hydrogen [8].

The production of green ammonia, synthesized through water electrolysis powered by renewable energy sources like solar photovoltaic (PV) systems, offers a climate-friendly alternative to conventional ammonia production methods. Over the past decade, solar PV technology has seen an 80% reduction in the Levelized Cost of Energy (LCOE), making it more cost-competitive than fossil-based electricity production by 2021 [11]. Onshore wind energy is also expected to become more economical than fossil fuels, and the use of Floating Photovoltaic Panels (FPV) provides an innovative and land-efficient solution for solar energy deployment [12].

In addition to conventional studies on ammonia production, previous research has analysed the energy and economic aspects of grey ammonia produced from natural gas [13], as well as a comparative economic and environmental assessment of grey (natural gas), blue (with carbon capture), and green ammonia (from hydropower and geothermal) [14]. Further studies have focused on the economic feasibility of green ammonia produced through water electrolysis powered by solar PV, wind energy, or grid electricity [15], while others have explored the environmental impacts of ammonia production using life cycle assessments (LCA) [16].

The transition to green ammonia, produced via water electrolysis powered by renewable energy, presents a sustainable future pathway. However, the adoption of this technology faces significant barriers, including high capital costs associated with electrolyzers and the required renewable energy infrastructure. Furthermore, there are technical and economic challenges that hinder large-scale implementation. Overcoming these barriers will require strong policy support and innovation to make green ammonia competitive with conventional methods like steam methane reforming [14].

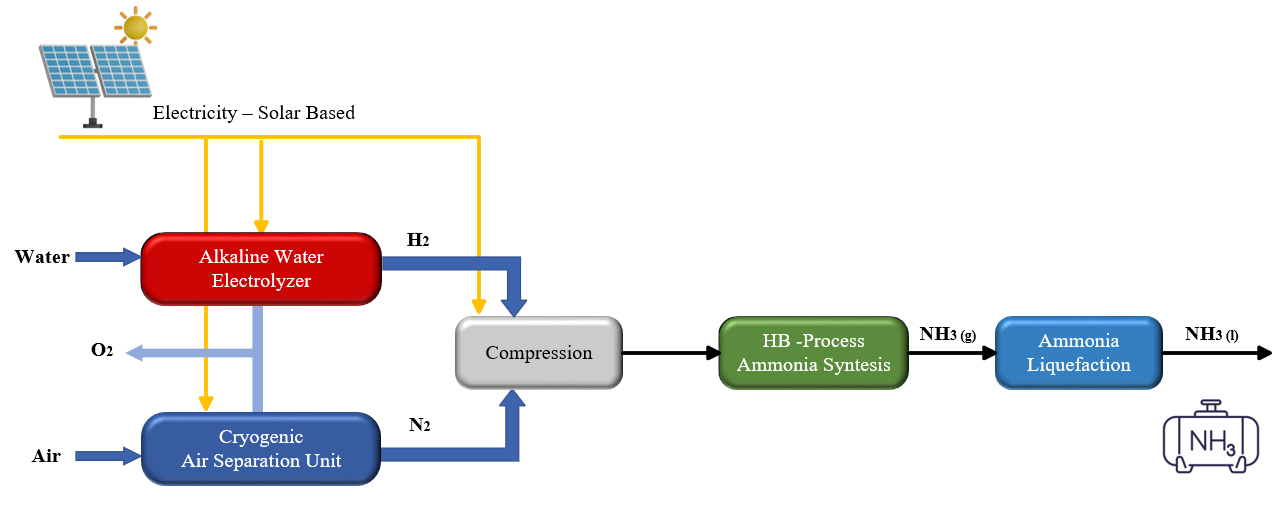
Recent advancements in green ammonia production have focused on improving the efficiency and cost-effectiveness of the process. One notable development is the integration of LNG cold energy, which can be used to enhance energy efficiency and reduce operational costs in ammonia production [17]. By utilizing the cryogenic energy from LNG regasification, processes such as air separation and ammonia liquefaction—both of which are energy-intensive—can be made more efficient, further improving the viability of green ammonia as a sustainable energy source. An added benefit of green ammonia production is the generation of high-purity oxygen as a byproduct of the water electrolysis process. This byproduct has the potential to provide additional revenue streams, which can offset the production costs of green ammonia, improving its overall economic feasibility [14].

The objective of this research paper is to explore the integration of LNG cold energy into green ammonia production. The proposed integration model combines Alkaline Water Electrolysis (AWE), Cryogenic Air Separation (CASE), and Ammonia Synthesis (AS) units. This study will simulate and analyze the energy and exergy efficiencies, exergy destruction profiles, and economic benefits of utilizing Floating Photovoltaic (FPV) systems located near LNG regasification terminals to produce green ammonia. Additionally, the high-purity oxygen produced as a byproduct of the water electrolysis process will be evaluated for its potential to provide additional economic value. This oxygen can be sold as a side product, which could offset production costs, improve the financial feasibility of green ammonia, and enhance the overall economics of the process.

# EXPERIMENTAL

## Process Description

The production scheme of green ammonia in Fig.1 is primarily based on the input feed of water, which undergoes alkaline water electrolysis (AWE) to yield hydrogen gas (H2), known as green hydrogen, and subsequently liquid ammonia. The total energy requirements of the entire process, particularly electrolysis, serve as the basis for determining the need for solar panels. Additionally, nitrogen gas (N2) is generated through the cryogenic air separation unit process (CASU). These two compounds are then reacted in the reactor for the ammonia synthesis process. Given the typical scale of an ammonia facility in Indonesia, the design basis has been chosen with a production capacity of 2,000 tNH₃/day, equivalent to 730,000 tNH₃/year [18].



**FIGURE 1**. Block flow diagram of proposed system

The AWE mainly produced renewable hydrogen, it is the most established technology and doesn’t require precious metals as catalysts with features a moderate efficiency of 59-70% [19]. Here, pure water with flow rate of 7365.81 kmol/hr operated in 80°C of temperature at 5 bar which is typically operated at approximately 70-90°C at a maximum pressure of 30 bar [3]. In this simulation, the LHV efficiency of electrolyzer 73% was selected. It needs 660 MW power to get 7365.81 kmol/h pure hydrogen gas with 99.99% purities. The reaction occurs as follows:

|  |  |  |
| --- | --- | --- |
| Anode: |  | (1) |
| Cathode: |  | (2) |
| Overall: |  | (3) |

The feed to the Cryogenic Air Separation Unit (CASU) is ambient air with a flow rate of 3120 kmol/hr, consisting of 79% N2 and 21% O2. The air is compressed to 8.23 bar at 30°C using a two-stage compressor, with LNG acting as the cooling medium. LNG is composed of 91.8% methane and 0.06% ethane, and is pumped to 40 bar before being expanded for the regasification process.The compressed air is cooled to -167°C using cold energy from LNG in the Air Separation Unit (ASU), before entering the flash separator. The liquid phase from the separator enters the Low-Pressure Column (LPC), and the vapor phase moves to the High-Pressure Column (HPC). The columns separate oxygen and nitrogen, with nitrogen being the lighter component. After distillation, the nitrogen is blended with hydrogen from the electrolyzer, while oxygen is mixed with oxygen gas produced in the electrolyzer for sale as a side product.

|  |  |  |
| --- | --- | --- |
|  |  | (4) |

The hydrogen-nitrogen mixture is compressed to 150 bar at 170°C using a multistage compressor. This compressed gas enters the ammonia converter, where ammonia is synthesized at 150 bar and 397°C, following the reaction in Gibbs reactor. The product is then cooled to 40°C with cooling water, followed by further cooling to -5°C using LNG cold energy. The mixture is then fed into a flash drum operating at 5 bar and -39°C. The overhead product is recycled, while the bottom product is liquid ammonia at -39°C and 5 bar pressure, with a purity of 99.96% by mass and a flow rate of 4917 kmol/hr. Shortly, LNG serves as the cooling agent in both the Air Separation Unit (ASU) and the ammonia liquefaction process.

## Simulation

The simulation schematic of the process is shown in Fig. 2. This simulation is performed using Aspen Plus V14. For hydrogen production, the ELECNRTL thermodynamic model is used, which is well-suited for handling electrolyte systems like alkaline water electrolysis. It accurately represents phase equilibria and interactions in aqueous systems [20]. For ammonia synthesis, the Peng-Robinson equation of state (EOS) is applied, as it is ideal for non-electrolyte systems involving gases like hydrogen and nitrogen and is widely used for modeling phase behavior in ammonia synthesis [21]. The overall process benefits from using ELECNRTL for electrolyte sections and Peng-Robinson for gas-phase modeling. The important compositions in the simulation are presented in Table 1, and the assumptions for the simulation are shown in Table 2.

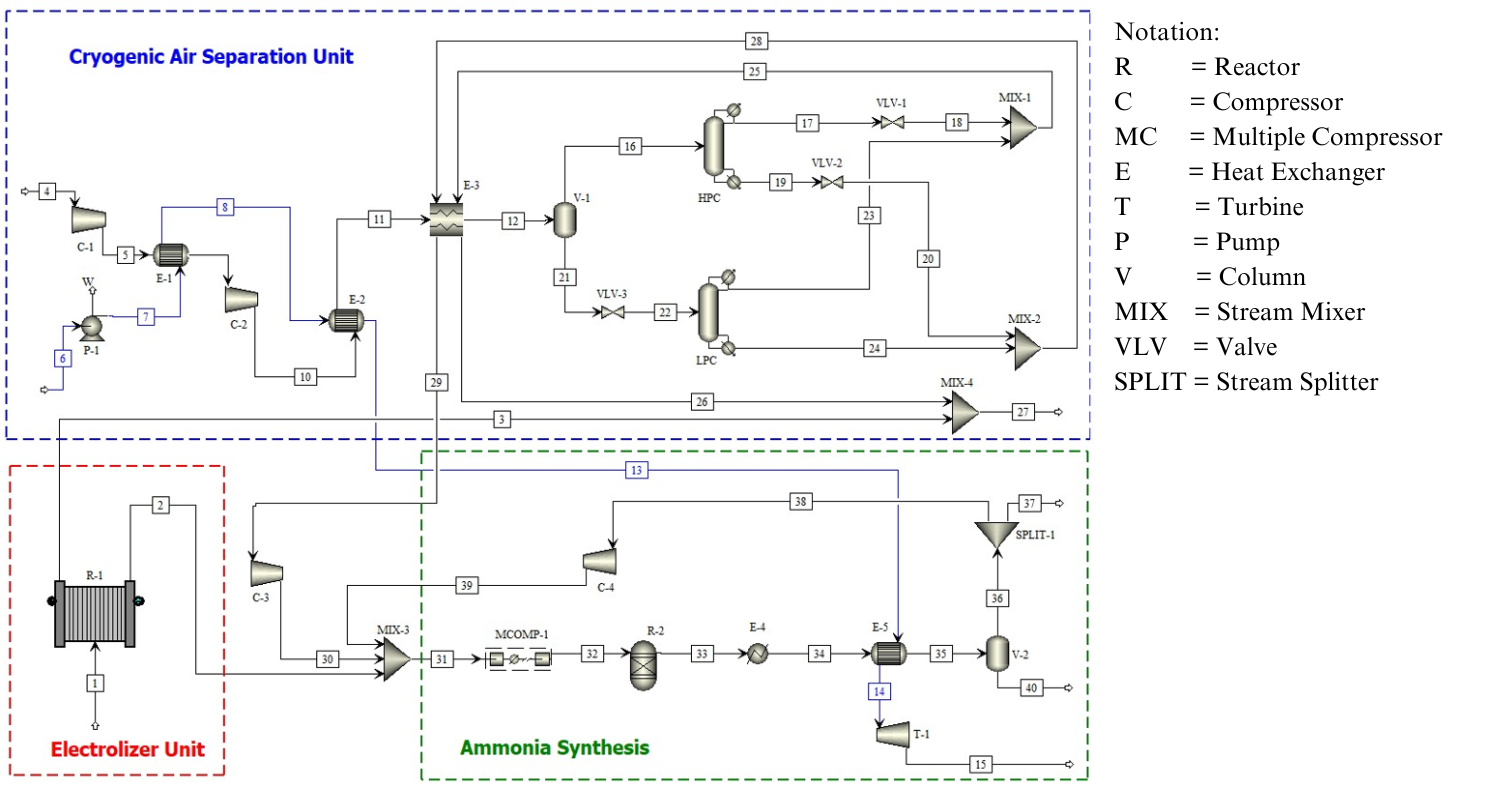
The Alkaline Water Electrolyzer (AWE) is modeled using an electrolyzer reactor R-1 within the scope module. The AWE configuration is designed to achieve a hydrogen production rate of 7365.81 kmol/h, with calculations revealing a power input of 660 MW and efficiency values: HHV efficiency of 86%, LHV efficiency of 73%, and voltage efficiency of 85%. To achieve this power level, 8000 stacks are used, with each stack containing 3500 cells. Consequently, the cell current, voltage, and power are calculated as 13.68 A, 1.72 V, and 23.57 Watts, respectively. Additionally, the water consumption rate required is 7149 kmol/h, and the side product oxygen has a purity of 99.99%, amounting to 3575 kmol/h.

On the Cryogenic Air Separation Unit (CASU), air enters through filters and is compressed up to 8.22 bar by compressors C-1 and C-2, with intercoolers E-1 and E-2 using LNG as a cooling medium. Subsequently, the air is cooled to -167°C using a cooling medium from the recycle stream of HPC and LPC. In the HPC, the reflux ratio is 1.2 with a condenser pressure of 7.6 bar and a reboiler pressure of 7.8 bar, with N2 as the light key. Similarly, in the LPC, the reflux ratio is 1 with a condenser pressure of 5 bar and a reboiler pressure of 5.1 bar. Both columns are simulated using DSTWU, a shortcut distillation model used to estimate reflux ratios, or the number of stages required to achieve the desired separation results.

|  |  |  |  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| **Table 1**. Composition of important streams | | | | | | | | | | |
|  | **Water** | **2** | **3** | **Air** | **16** | **17** | **19** | **21** | **23** | **24** |
| Components | Mol Fraction | | | |  |  |  |  |  |  |
| H2O | 1.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| O2 | 0.0000 | 0.0001 | 0.9999 | 0.2100 | 0.1478 | 0.0002 | 0.9943 | 0.2970 | 0.0004 | 0.9976 |
| H2 | 0.0000 | 0.9999 | 0.0001 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| CO2 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| NH3 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |
| N2 | 0.0000 | 0.0000 | 0.0000 | 0.7900 | 0.8522 | 0.9998 | 0.0057 | 0.7030 | 0.9996 | 0.0024 |
|  | **25** | **27** | **28** | **30** | **31** | **33** | **35** | **36** | **40** |  |
| Components | Mol Fraction | | | |  |  |  |  |  |  |
| H2O | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | 0.0006 | 0.0000 | 0.0006 |  |
| O2 | 0.0027 | 0.9994 | 0.9962 | 0.0003 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |  |
| H2 | 0.0000 | 0.0000 | 0.0001 | 0.7493 | 0.7792 | 0.6682 | 0.6009 | 0.6682 | 0.0000 |  |
| CO2 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 |  |
| NH3 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.1133 | 0.2846 | 0.3119 | 0.2846 | 0.9968 |  |
| N2 | 0.9997 | 0.0006 | 0.0038 | 0.9997 | 0.1075 | 0.0471 | 0.0871 | 0.0472 | 0.0026 |  |

**Table 2**. Provides all the following assumptions of simulation.

|  |  |
| --- | --- |
| **Process** | **Assumption** |
| Overall Process | The equations are applied under steady-state conditions |
| Turbines, pumps, and compressors operate adiabatically and isentropic efficiency of 80% |
| All separators are adiabatic and complete |
| AWE | KOH used as an electrolyte with 30 wt.% [22] |
| AS Unit | Ammonia purge for recycle was 0.5% |



**FIGURE 2**. Simulation schematics of proposed system

Pure nitrogen from CASU is compressed to 25 bar and then mixed with hydrogen from an electrolyzer before being further compressed to 150 bars using the multiple compressor MCOMP-1, with its intercooler using cooling water. The mixed gas enters the adiabatic reactor Gibbs R-2 at 150 bar pressure. The resulting gas from the reaction is cooled by cooler E-3 using cooling water utility, then further cooled to -5°C to obtain liquid ammonia in heat exchanger E-4, which utilizes LNG and serves as the LNG regasification process to convert LNG into natural gas. Subsequently, the liquid ammonia is separated in flash separator V-2. The gas exiting is recycled, with 0.5% purge gas. More detailed operational parameters for each equipment are provided in Table 5.

**Table 3**. Thermodynamic conditions of all process streams.

|  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- |
| **Stream** | **T (°C)** | **P (kPa)** | **F (kmol/h)** | **Stream** | **T (°C)** | **P (kPa)** | **F (kmol/h)** |
| 1 (Water) | 80.00 | 2600.00 | 7149.16 | 21 | -169.00 | 800 | 1250.627 |
| 2 | 80.00 | 2500.00 | 7149.52 | 22 | -174.53 | 540 | 1250.627 |
| 3 | 80.00 | 2500.00 | 3574.22 | 23 | -179.14 | 500 | 878.7098 |
| 4 (Air) | 20.00 | 101.33 | 3000.00 | 24 | -164.11 | 510 | 371.9175 |
| 5 | 141.77 | 293.84 | 3000.00 | 25 | -179.14 | 500 | 2368.26 |
| 6 (LNG) | -162.00 | 101.33 | 9213.70 | 26 | -128.09 | 510 | 631.74 |
| 7 | -156.11 | 400.00 | 9213.70 | 27 | 45.53 | 510 | 4205.964 |
| 8 | -141.45 | 400.00 | 9213.70 | 28 | -164.14 | 510 | 631.74 |
| 9 | 25.00 | 293.84 | 3000.00 | 29 | -128.09 | 500 | 2368.26 |
| 10 | 144.19 | 822.76 | 3000.00 | 30 | -11.17 | 2500 | 2368.26 |
| 11 | 30.00 | 822.76 | 3000.00 | 31 | 108.05 | 2500 | 35214.03 |
| 12 | -167.00 | 822.76 | 3000.00 | 32 | 170.00 | 15000 | 35214.03 |
| 13 | -140.76 | 400.00 | 9213.70 | 33 | 397.01 | 15000 | 30516.72 |
| 14 | 24.73 | 400.00 | 9213.70 | 34 | 40.00 | 14995 | 30517.16 |
| 15 | -45.69 | 100.00 | 9213.70 | 35 | -5.00 | 14990 | 30517.16 |
| 16 | -169.00 | 800.00 | 1749.37 | 36 | -39.00 | 400 | 25825.37 |
| 17 | -173.48 | 760.00 | 1489.55 | 37 | -39.00 | 400 | 129.1269 |
| 18 | -179,15 | 500.00 | 1489.55 | 38 | -39.00 | 400 | 25696.25 |
| 19 | -157,77 | 780.00 | 259.82 | 39 | 126.36 | 2500 | 25696.25 |
| 20 | -163,89 | 520.00 | 259.82 | 40 | -39.00 | 400 | 4691.789 |

**Table 4**. Specification of the equipment employed in the proposed process.

|  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- |
| **Compressor** | | | | | |
| Unit | Adiabatic efficiency (%) | Power (kW) | Outlet pressure (bar) | P ratio | ∆P (bar) |
| C-1 | 0,8 | 2971,2 | 2,94 | 2,9 | 1,93 |
| C-2 | 0,85 | 2906,6 | 8,23 | 2,8 | 5,29 |
| C-3 | 0,72 | 2185,0 | 25 | 5 | 20 |
| C-4 | 0,8 | 39181,4 | 25 | 6,25 | 21 |
| **Multistage Compressor** | | | | | |
| Unit | Adiabatic efficiency (%) | Power (kW) | Outlet pressure (bar) | Cooling duty (kW) | No of stages |
| MCOMP-1 | 0,8 | 5929,361 | 8,23 | -5724,58735 | 2 |
| **Alkaline Water Electrolyzer (AWE)** | | | | | |
| Unit | Temperature (°C) | Power (kW) | HHV Efficiency (%) | Spesific Energy (kJ/kgH2) | Pressure (bar) |
| R-1 | 80 | 488328 | 0,86 | 123158 | 1 |
| **Ammonia Converter** | | | | | |
| Unit | Tin (°C) | Duty (kW) | Pressure (bar) |  |  |
| R-2 | 170 | 0 | 150 |  |  |
| **Cooler dan Heater** | | | | | |
| Unit | T in (°C) | Duty (kW) | T out (°C) |  |  |
| E-3 | 401,692 | -127423 | 40 |  |  |
| **Heat Exchanger** | | | | | |
| Unit | LMTD (°C) | Duty (kW) | UA (cal/sec.K) | Exchanger Area (m2) | ∆P (kPa) |
| E-1 | 227,95 | 2860 | 2997 | 14 | 0 |
| E-2 | 223,29 | 2812 | 3008 | 14,8 | 0 |
| E-4 | 80,53 | 31364 | 93029 | 458,23 | 0 |
| **Column** | | | | | |
| Unit | Reflux ratio | No of Stages | Qr (MW) | Qc (MW) | ∆P (bar) |
| HPC | 1,2 | 34 | 1,7964 | -3,9198 | 0,2 |
| LPC | 1 | 45 | 2,0484 | -2,2350 | 0,1 |
| **Flash Separator** | | | | | |
| Unit | T(°C) | Duty (kW) | Pressure (bar) |  |  |
| V-1 | -169 | -1589,4 | 8 |  |  |
| V-2 | -40 | 11563 | 4 |  |  |

# process analysis

## Exergy Analysis

One of the most crucial analyses for assessing the thermodynamic performance of a process is exergy analysis, which is discussed in this section. The primary objective is to quantify the destructed exergy, determine the contribution of each unit and piece of equipment to exergy destruction, calculate the total exergy efficiency, and compare these results with other studies. Generally, the exergy of a stream is the sum of its physical and chemical exergies, as described below [23].

(4)

(5)

(6)

Where and represent the specific enthalpy and entropy of the stream at ambient temperature and pressure, respectively. In this analysis, the calculation of physical exergy is performed using Aspen Plus, followed using Equation 6 to compute chemical exergy. The exergy destructions of the equipment are determined using Equation 7, with the results presented in Table 5. The exergy efficiency of the entire process is evaluated using Equation 8 [24].

(7)

(8)

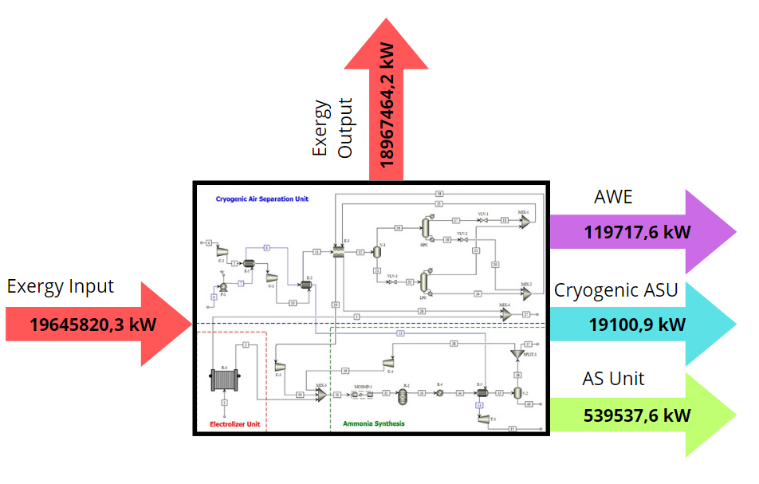
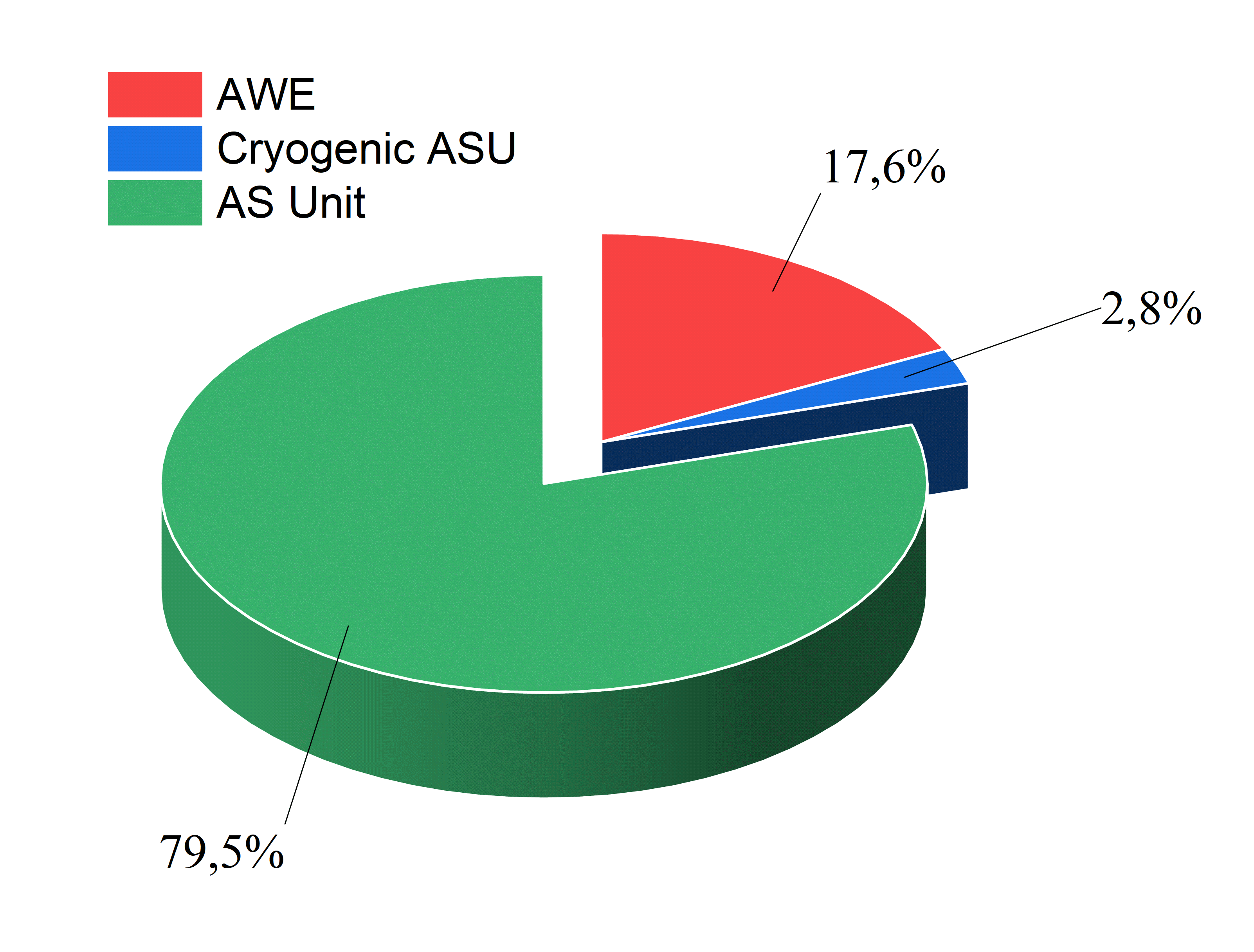
**Table 5**. Equations exergy destruction of components and the whole process.

| **Component** | **Exergy In** | **Exergy Out** |
| --- | --- | --- |
| Compressor [25] |  |  |
| C-1 |  |  |
| C-2 |  |  |
| C-3 |  |  |
| C-4 |  |  |
| MC-1 |  |  |
| Pump [25] |  |  |
| P-1 |  |  |
| Flash drum [26] |  |  |
| V-1 |  |  |
| V-2 |  |  |
| Turbine [27] |  |  |
| T-1 |  |  |
| Reactor [28] |  |  |
| R-1 |  |  |
| R-2 |  |  |
| Heat Exchanger [27] |  |  |
| E-1 |  |  |
| E-2 |  |  |
| E-3 |  |  |
| E-5 |  |  |
| Column [29] |  |  |
| HPC |  |  |
| LPC |  |  |

To gain a comprehensive understanding of the exergy flow, a simplified Sankey diagram of the proposed process is illustrated in Fig.3. This diagram clearly depicts the exergy flow within the process and the exergy loss associated with each unit. As shown in Fig.3, the system receives a total exergy flow of 19,630,548 kW, of which only 18,952,193 kW is recovered. Furthermore, the exergy destruction for the AWE, Cryogenic ASU, and AS Unit is 119,718 kW, 19,100 kW, and 5,39,538 kW, respectively. The AWE unit itself comprise only one component that is electrolyser while Cryogenic ASU and AS Unit have various components. According to Fig.3, the contributions to exergy destruction from the AWE, Cryogenic ASU, and AS Unit subsystems are 17.6%, 2.8%, and 79.5%, respectively. From that figure, it indicates that within the AS Unit, the Ammonia converter (R-2) and point of feed ammonia reactor mixture (MIX-3) contribute 81.7% to the total exergy destruction. Therefore, it is evident that the AS Unit has the highest exergy destruction within the proposed process, and within the AS Unit, the Cryogenic ASU has the lowest exergy destruction.

Fig.4 illustrates the contributions to exergy destruction by the equipment within the Cryogenic ASU and AS Unit separately. In the Cryogenic ASU unit, the heat exchanger that integrate subsystem with LNG as cold stream has the largest contribution to exergy destruction at 6848 kW (35.85%), 3926 kW (20.56%), and 4092 kW (21.42%) for multiple heat exchanger MHE1, heat exchanger E-1, and E-2, respectively. In the AS Unit, ammonia converter R-2 contribute the largest exergy destruction around 56.62% or 45.01% of overall proposed system. This occurs because the chemical exergy of the components within the AS Unit has high values, due to the presence of a recycle stream with the purge gas assumed to be 0.5%. The second largest exergy destruction at AS Unit is MIX-3 which contribute 25.08% or 19.95% of the total process exergy destruction. This phenomenon occurs because, when two gases at different temperatures are mixed, energy must be redistributed to reach thermal equilibrium. This process is not completely reversible (it cannot be reversed without incurring energy loss) and results in increased entropy, thereby causing exergy loss.

The overall exergy efficiency of 96.54% (Equation 8) underscores the thermodynamic advantages of the proposed process, minimizing energy losses and optimizing resource utilization. Notably, the exergy destruction in this process is comparable to that in Scenario 1, with a key distinction being the use of LNG cold energy in the CASU and ammonia liquefaction. Exergy destruction arises primarily from temperature differences during heat transfer, fluid friction within heat exchanger pipes, and exergy loss to the surroundings, with significant contributions from E-1, E-2, and MHE1 in the CASU. Mitigation strategies include reducing temperature differences between fluids [30], minimizing pressure drops, optimizing flow rates, enhancing geometric designs, and employing advanced techniques like nanofluids and multi-objective optimization to improve efficiency and reduce thermodynamic irreversibilities [31].



**FIGURE 3**. Contribution of different units of the proposed process to exergy destruction

A pie chart with numbers and percentages

Description automatically generated A pie chart with numbers and a number of different colors

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**FIGURE 4**. Contributions of equipment in the Cryogenic ASU and Ammonia Synthesis Unit to exergy destruction

## Energy Analysis

The energy efficiency of the overall process and its subsystems is discussed. To calculate the total energy efficiency of the process, Equation (9) [24] is used. This equation also provides the basis for formulating the energy efficiencies of all individual units. Thus, the overall energy efficiency for the proposed system is represented by Equation (10).

|  |  |  |  |
| --- | --- | --- | --- |
|  |  | | (9) |
|  |  | (10) | |

In the above equations, HHV denotes the higher heating value in MJ/kg, with ammonia having a value of 22 MJ/kg (6.1045 kW/kg) [32], denotes the produced thermal energy, represents the net produced power for example from turbine T-1, and is the power consumed in compressors and pumps. In Equation (10), is the difference between the total power consumption of AWE, compressors, and pumps, and the amount of power produced by the turbines of the AS Unit. Additionally, the value of equals the difference between the total heat recovered in reactor R-1 and the required heat in the reboiler of the distillation columns HPC and LPC.

A graph of energy and energy production

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**FIGURE 5**. Energy and exergy efficiencies of various ammonia production technologies. (A) Power-to-ammonia [33] (B) Ammonia from hydropower [34], (C) Ammonia from nuclear [34], (D) Ammonia from biomass [34], Solar photovoltaic-based ammonia production [35], (F) This study

From Equation (10), the energy output from the ammonia product is calculated as 481.945 MW, with an energy input of 872.39 MW, resulting in an overall system energy efficiency of 55.25%. This demonstrates that the proposed system optimizes energy usage, offering competitive performance for real-world applications. Additionally, the high efficiency suggests potential for reduced costs and improved sustainability. The energy efficiencies of different units in the process are compared in Fig.5.

## Economic Analysis

The Levelized Cost of Ammonia (LCOA) is selected as the primary metric for evaluating and comparing the economic performance of various ammonia production methods. The LCOA accounts for both the capital and operational costs required per unit of ammonia produced. By incorporating these factors, the LCOA provides a comprehensive measure of the economic viability of different ammonia production technologies, allowing for a clearer comparison of their cost-effectiveness and efficiency.

The LCOA is calculated using Equation (12), with the components of the Total Operating Cost (TOC) detailed in Table 6. This calculation follows the techno-economic analysis conducted by Tjahjono et al. [14], ensuring that the results are directly comparable with their study. This methodology provides a standardized approach to assessing the economic feasibility of ammonia production technologies, facilitating a consistent comparison across different studies and production methods. However, one significant difference is the inclusion of revenue from the LNG regasification credit, which is adjusted to $2.5/MMBTU, in line with typical LNG regasification terminals in Indonesia. In this economic analysis, it is also assumed that the additional TCR resulting from the installation of cryogenic piping and heat exchangers is borne by the regasification party. Despite this, the integration of the system still offers economic benefits for both the ammonia producer and the regasification terminal owner.

**Table 6**. Economic Assumptions and Equations

| **Components** | **Equation** |
| --- | --- |
| Annual plant capacity | 730000 TPY [18] |
| Capacity Factor | 90% |
| Discount Rate | 8% |
| LNG Regasification Credit | 2.5$/MMBTU [36] |
| Solar Grid Electricity Cost | 45/MWh |
| **Major Installed Cost** | |
| Mechanical Vapor Compression (MVC) [37] |  |
| Electrolyser Modules (EM) [37] |  |
| Air separation Unit (ASU) [38] |  |
| Haber-Bosch Reactor (HB) [38] |  |
| Purchased Equipment Cost (PEC) [38] |  |
| Utilities and Balance of Plant (BoP) [38] |  |
| **Total Cost Requirement [30]** | |
| Total Plant Cost (TPC) |  |
| Spare Parts Cost (SPC) |  |
| Start-up Cost (SUC) |  |
| Owner’s Cost (OC) |  |
| Interest during Construction (IDC) |  |
| Working Capital (WC) |  |
| Total Capital Cost Requirement (TCR) | TCR = TPC +SPC +SUC+OC+IDC+WC |
| **Fixed Operational Cost** | |
| Direct Labor (DL) [39] |  |
| Maintenance (M) [39] |  |
| Insurance and Local Taxes (ILT) [40] |  |
| Administration and General Overheads (AGO) [39] |  |
| Amortization (A) [41] | , where alkaline electrolizer has EML 10 years |
| Fixed Operation Cost (FOC) | FOC = DL + M + AGO + ILT +A |

Based on the LCOA parameter and compared to other ammonia production technologies illustrated in Fig.6, ammonia production using AWE with solar photovoltaic as the electricity source and integrated with LNG terminal regasification is $469/t-NH3, whereas study from Tjahjono et al reveal thatgrey ammonia through SMR of methane is $297/t-NH3. In contrast, the cost of ammonia production using AWE powered by geothermal and hydropower electricity is $835/t- NH3 and $696/t- NH3, respectively. In this study, when the same technology is used without integration with LNG cold energy, the cost is $765/t- NH3. Additionally, the study indicates that the pure oxygen produced at a rate of approximately 3,394.74 tpd could potentially be sold for $140/t-O2, resulting in a more competitive LCOA of around $250/t- NH3. Therefore, it can be concluded that integrating LNG cold energy into green ammonia production offers a competitive economic advantage comparable to the currently mature SMR technology.

A graph of different colored bars

Description automatically generated

**FIGURE 6**. LCOA comparison of various ammonia production technologies. (a) grey ammonia from SMR, (b) blue ammonia from scenario a + CCS, (c) green ammonia AWE-hydropower, (d) green ammonia AWE-geothermal [14], (e) this study without LNG integration, (f) this study

## Environmental Analysis

Green ammonia production is generally associated with zero emissions. However, a detailed examination of the environmental impact of various ammonia production technologies has been conducted by Singh et al. [16] using the Life Cycle Assessment (LCA) method. Their findings indicate that green ammonia produced through Alkaline Water Electrolysis (AWE) and powered by solar PV electricity exhibits a climate change indicator value of 1.2775 kgCO2-eq. This is 57.8% lower compared to the conventional Steam Methane Reforming (SMR) technology. Nonetheless, the PV electrolysis device presents concerns due to its high values for acidification, eco-toxicity, and human toxicity. For instance, the human toxicity value is 1.366 kg 1,4-DB-eq, and the freshwater aquatic eco-toxicity is 0.753 kg 1,4-DB-eq per kg of ammonia produced. These values are considerably high compared to all other resources investigated. This elevated environmental impact arises from the hazardous materials used in the manufacturing of PV cells and the chemicals used for cleaning, including hydrochloric acid, sulfuric acid, nitric acid, hydrogen fluoride, 1,1,1-trichloroethane, and acetone.

The integration of green ammonia production with LNG regasification offers economic advantages through the utilization of cold energy. While the green ammonia production process itself achieves zero direct CO2 emissions, the indirect emissions must also be considered. The regasified LNG, amounting to 185 MMSCFD, has the potential to generate CO2 emissions of approximately 9,847.86 tCO2-eq per day, assuming it is used as fuel and follows the complete combustion reaction of methane. This significant level of indirect emissions highlights the necessity for further environmental analysis to assess the impact of integrating green ammonia production with LNG regasification.

Given the potential for significant CO2 emissions and the environmental concerns associated with PV electrolysis devices, a comprehensive environmental analysis is essential. This analysis should evaluate the overall environmental impact of integrating green ammonia production with LNG regasification. The goal is to ensure that the benefits of using green ammonia do not lead to unintended negative environmental consequences, thus promoting a truly sustainable and eco-friendly energy solution.

# CONCLUSION

Based on the analysis of the green ammonia production scheme with the integration of LNG cold energy using the 4E method, the main conclusions of the analysis are as follows:

* The energy and exergy efficiencies are 55.24% and 96.54%, respectively.
* From an exergetic perspective, the total exergy destruction amounted to 678,355 kW. The Air Separation Unit (ASU) subsystem contributed the largest portion of this destruction, accounting for 79.54%, followed by the Alkaline Water Electrolysis (AWE) subsystem at 17.6%, and the Carbon Air Separation Unit (CASU) at 2.8%. Moreover, among the various equipment, the ammonia converter exhibited the highest share of exergy destruction
* The proposed system has zero emissions of CO2, but it potentially produces 9,847.86 tCO2-eq per day from methane combustion of natural gas that is available from LNG regasification.
* According to economic analysis, the LCOA of the proposed system was 469 $/t-NH3 that is more competitive compared to the same system without LNG integration.

While promising, the study's reliance on specific assumptions and exclusion of upstream emissions or the full lifecycle impact of LNG regasification limits its scope. Pilot-scale studies and lifecycle assessments are needed to validate its applicability. Future work should optimize subsystem designs, explore cleaner energy sources to reduce indirect CO2 emissions, and enhance process control. These steps will support the scale-up and practical implementation of sustainable ammonia production.

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