

# Technological Assessment of Commercial Ammonia Synthesis Methods in Coastal Areas of Germany

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**ABSTRACT** – Ammonia synthesis through renewable energy technology can transform the energy and fertilizer markets of the future. The study covers the techno-economic environmental analysis of conventional ammonia synthesis via fossil fuels and green ammonia synthesis via renewable energy technology in coastal areas of Germany. The basis for comparison to investigate the economic feasibility of different ammonia plants is levelized cost of ammonia production. The conventional ammonia synthesis plant uses Steam Methane Reforming and Haber Bosch Process plants that requires natural gas as a source of hydrogen and air as a source of nitrogen. Similarly, the green ammonia plant uses water electrolysers (Alkaline/PEM electrolysers), Air Separating Unit, Water Desalination and Haber Bosch Process plants that are all powered by onshore wind farm

## 1. INTRODUCTION

Ammonia is one of the commonly synthesized chemicals with a global production of 180 million tons (2017) [1] [2]. With a growth rate of around 2% annually for the last two decades, Ammonia is projected to reach 360 million tons in 2030 [3]. Ammonia produced is majorly consumed by fertilizer industry. Fertilizer industry consumes around 80% of global ammonia production [4].

However, the process of ammonia production is dependent entirely on fossil fuels for energy requirements and for supplying hydrogen. Approximately 72% of total ammonia production is based on natural gas through steam methane reforming (SMR) in 2018. Moreover, approximately all of the hydrogen, required to feed ammonia synthesis loop, is produced from fossil fuels and majorly from natural gas [5]. The production of ammonia consumes about 1.8 – 3.0 % of global energy production and is responsible for emitting 1% of carbon emissions (around 290 mega tons) annually around the globe [6]. Natural gas release around 2.7 metric tons, while coal release 3.4 tons of greenhouse gases (GHG) for every ton of ammonia produced [7].

With the current practices and predicated increase in ammonia production, will increase carbon emissions and the growing levels of carbon emissions, also called as greenhouse gases (GHG), will led to further increase in increase global temperatures (global warming). This has been studied and proved [8]. Moreover, the prices of natural gas are expected to increase in the future and increase in carbon tax in developed nations may disrupt the fertilizer industry in the future [9] [10]. This could also adversely impact world population.

So, in order increase the sustainability of ammonia production process, the process can be decarbonised by incorporating renewable energy technology. The renewable energy

technology like solar, wind can provide energy for air separation, water electrolysis and ammonia synthesis loop [9]. This concept is not new and has been investigated previously in the past.

An example is that of Morgen et. al [11] investigated, in 2014, the feasibility producing ammonia from diesel generator and wind power with a back-up diesel generator for a remote island of USA called Monhegan. The energy produced from the wind farm was used to run cryogenic air distillation, alkaline electrolysers for electrolysis of water, mechanical vapour compression for desalination of water and Haber-Bosch (HBP) synthesis loop. The levelized cost of ammonia (LCOA) was estimated to be \$1224/ton of ammonia.

Similarly, Nayak-Luke et. al [12] investigated, in 2018, the feasibility of producing ammonia from solar and wind farm for a remote island of Scotland, Lerwick. Using estimates from 2025 and 2030, LCOA was estimated at £588 per ton of ammonia, with a levelized cost of electricity (LCOE) at £45.7 per MWh, renewable energy mix of 90% wind and 10% solar technology, electrolyser capital at £308 per kW and ASU/HBP process minimum power is taken 20% of rated power.

The objective of this work is to study the potential of producing ammonia in a sustainable way. Techno-economic studies compares the existing SMR – HBP with green ammonia production using alkaline electrolyser technology and PEM electrolyser technology. Using DWSIM, a steady-state chemical process simulator, ASU, SMR, and HBP are simulated for small scale ammonia (300 tons per day of ammonia) [13]. The small scale is considered in this work as the electrolysis technology is developed but scale of manufacturing is undersized when compared large scale SMR-HBP plants [14].

The economics studies use capital costs and operating costs to calculate the levelized cost of ammonia (LCOA). LCOA will be a basis to compare conventional SMR-HBP with alkaline electrolyser-HBP and PEM electrolyser-HBP. Moreover, the economics study focuses on Germany as it is one of the leading European Union (EU) countries with renewable energy generation of around 42% [15].

## 2. LITERATURE REVIEW

### 2.1 AMMONIA SYNTHESIS

Ammonia is a vital raw material for fertilizer and agriculture industry, it is consumed around 80 % from total ammonia consumption in 2016 [4]. The consumption by weight of ammonia in fertilizer industry was 143 million tonnes in 2017 [2]. Apart from fertilizer synthesis, ammonia is used to manufacture nitric acid, which is post processed to

make explosives, fibres, plastics, dyes, pharmaceutical drugs and ammonium nitrate [4]. The other commercial uses of ammonia also include the production of chemicals for SCR systems, refrigeration units, wastewater treatment, metal treatment, leather, rubber, paper, household cleaning, food and beverage industries [16] [17].

The pathways for production of ammonia existing currently are depicted in figure 1 and compared in table 1. Although major production of ammonia is reliant on fossil fuels especially natural gas, in the past ammonia was produced by utilizing hydroelectric power in Rjukan (Norway), Glomjford (Norway), Aswan (Egypt), Reykjavik (Iceland), Cusco (Peru) [18].

However, currently there are no commercial plants operational due to two reasons. Firstly, the very high energy requirements to overcome the thermodynamic barrier, 39.4 kWh/kg of hydrogen (ideal), to produce hydrogen water. Secondly, the electrolysis technology is developed but scale of manufacturing is undersized when compared to commercial SMR-HBP producing 1000-1500 tonnes per day [19] [20].

The use of solar and wind energy technology is limited. However, the feasibility of using renewable energy technology for ammonia production has been studied recently by the demonstration plants in Oxford, UK and Fukushima, Japan. The former used wind energy and the latter used solar energy to obtain green hydrogen from water via electrolysis, nitrogen from air separation and to power the Haber Bosch process to make green ammonia [21] [22].

Moreover, Electrochemical ammonia synthesis is a relatively new process to ammonia directly through electrolytic cells [23]. It is an alternative to HBP that combines nitrogen from air with hydrogen or any hydrogen-containing compound (like water) in an electrolytic cell [24]. Using water as a hydrogen carrier, if commercial success is achieved, will save capital costs, energy consumption and reduce carbon footprint to produce hydrogen for ammonia synthesis [19]. However, this process is currently under research and development and there is no commercial system available in the market [24].

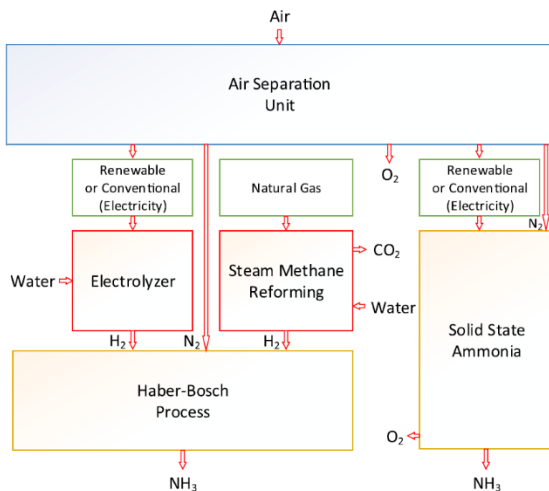


Figure 1: Ammonia production routes [5]

### 2.2.1 HABER BOSCH PROCESS (HBP)

HBP is the most common and commercial process for ammonia synthesis. Most of the ammonia is produced globally by utilization of HBP on an industrial scale. Although the HBP may have different configurations in every industry, the basic process flow and features are the same and well established. It

involves compressors, a reactor, a flash drum and a network of heat exchangers [19].

The feed gases, with a purity of 99.99% for both nitrogen and hydrogen, are compressed to the required pressure using intercooled staged compressors. The feed gases are heated to a required temperature and fed to catalyst beds, where nitrogen and hydrogen are reacted to form ammonia. The gases are then transferred to a condenser, where ammonia is separated from unreacted nitrogen and hydrogen. The unreacted hydrogen and nitrogen are compressed and recycled into the main feed stream. Moreover, to avoid the build-up of inert gases and having an adverse impact on efficiency, the gases are purged periodically [25] [19].

Table 1: Comparison of ammonia synthesis technologies [23] [17] [26] [27]

Production method	Energy consumption (kWh/ton NH3)	CO2 (kg/kg NH3)	Efficiency
SMR + Haber-Bosch	9,500	1.7-2.1	61-66%
Water Electrolysis + Haber-Bosch	12,000	-	Approx. 54%
SSAS	7,000 – 8,000	-	65-70%

The synthesis of ammonia is an exothermic reaction (-92 kJ/mol) involves mixing of hydrogen and nitrogen in a ratio of 3:1 as shown in the equation below. Additionally, the process includes iron-based catalyst with temperature and pressure ranging from 450–600 °C and 100–250 bar, respectively. The high operating parameters are a trade-off between the Le Chatelier's principle and the fast reaction rates. This converts 20-30 % of hydrogen into the final product, so in order to increase the yield to 98 % a recycle loop is used [5] [9].



## 3. METHODS

### 3.1 SYSTEM ECONOMICS

The equipment cost estimation methods are based on Morgan et. al [25], Turton et. al [28] and Feng et. al [29]. The approach is module costing technique which is used for making preliminary cost estimated for new chemical plants. The purchase cost of equipment is evaluated using for some base conditions (carbon steel equipment operating at ambient pressure) and then the cost of equipment is adjusted by multiplying factors if the equipment working at different conditions, other than base conditions. This usually depends upon the specific type of equipment, pressure and material of the equipment. Module costing technique will be applied for SMR, HBP and cryogenic ASU.

Similarly, the manufacturing/operating costs are based on approach used by Morgan et. al [25], Turton et. al [28] and Feng et. al [29]. This technique will be applied on SMR, HBP, cryogenic ASU, mechanical vapour compressor (MVC), alkaline electrolyser and PEM electrolyser process units.

Note: Prices calculated in this thesis are in US dollars in 2019.

### 3.2 HBP PRODUCTION LOOP

The scope of the work is the study of a 300 metric ton per day of ammonia. This will stoichiometrically require 246.7 tons of nitrogen and 53.3 tons of hydrogen per day as per equation 1. HBP loop as mentioned varies from industry to industry as per requirement of that specific plant. However, some equipment are common as per basic process flow. These include multi-staged compressor, recycle compressor, reactor, heat exchanger and separator. The actual HBP cycle of industrial scale is extensive design and beyond the scope of this thesis, so a preliminary analysis will be based on process simulations based on available literature.

The high purity nitrogen is fed from cryogenic air distillation, while high purity hydrogen is fed from either alkaline electrolyser plant, PEM electrolyser plant or SMR plant. After selecting the size and major components of the plant, HBP loop is developed on DWSIM simulation software (steady state). The pressure and temperature specifications of equipment are based on models by Morgen et. al [25] and Araújo et. al [30].

HBP converts high purity nitrogen and hydrogen feed to ammonia. The process takes place at high temperature (695 K) and pressure (156 bars). The process converts 26.4317% of hydrogen into ammonia, and the unreacted gases are recycled after separating ammonia.

Similarly, the equipment are sized and economic estimation is carried out using simulation and stream data from DWSIM flowsheet. The major equipment are compressors, heat exchangers, reactor and separator. Furnace is used in the case of SMR-HBP coupling, while in the case of alkaline electrolyser and PEM electrolyser, heat load is considered as electric load for simplification purposes.

Based on equipment sizing and stream data, the capital costs and manufacturing costs incurred in installing and operating HBP and EHBP (Electro Haber Bosch Process) loop with heat recovery unit (HBP-HR and EHBP-HR) are shown in table 2 and table 4 respectively.

Note: Streams labelled as ESTR-0XX and yellow in colour are energy streams while other streams blue in colour are material streams.

Table 2: Capital cost of installing HBP-HR and EHBP-HR loop

Equipment		Capital Cost (mil. \$)
Heat exchangers		11.31
Compressors		21.06
Reactors		2.46
Separator		2.55
Furnace		2.39
HBP-HR	C <sub>TM</sub>	46.92
	C <sub>GR</sub>	55.78
EHBP-HR	C <sub>TM</sub>	44.11
	C <sub>GR</sub>	52.11

#### 3.2.1 AMMONIA STORAGE SYSTEM

The storage tank are designed to store ammonia for about 30 days [31], so storage capacity required for a 30-day period

for 300 tonnes per day plant requires a tank of 9000 tonnes. Since it is a large volume, the capital investment cost for ambient pressure storage are lower than pressure storage [16]. The design pressure is in the range of 1.1 - 1.5 bars, including the static pressure of stored ammonia [16].

The amount of volume corresponding to a 30-day storage is approximately 13,184 m<sup>3</sup> and adding a 10% freeboard makes the final volume of storage around 14,503 m<sup>3</sup>. The optimum height to diameter ratio, to avoid wind overturning and stable for seismic loading, is 0.6 [32]. The diameter and the height of the tank are approximately 31m and 19m respectively. The inner tank is usually surrounded by an outer tank to contain leaked ammonia vapor and liquid [33] [34]. The outer tank is located at a maximum distance of 1.5m from inner tank minimum distance. The height and diameter of the outer tank is 20.5m and 32.5m respectively.

The inner tank stores refrigerated ammonia at -33°C, the refrigeration is provided by recompressing the boil-off from the tank [16]. The recompression is provided by 2 stage reciprocating compressors [16]. The refrigeration provided is depended on the boil-off from the tank, which in turn is depended on the heat flux of the inner tank and latent heat of vaporisation of ammonia [25] [35]. The tank is usually designed to keep the boil-off to be less than 0.04% of the ammonia stored in the tank [25] [35]. Assuming the boil-off of 0.04%, the amount of ammonia vapours entering the refrigeration unit is 3.6 tons per day for a fully filled ammonia tank. The refrigeration loop is developed in DWSIM software as per the approach by Morgan et. al [25], Belapurkar et. al [35] and Webb et. al [36].

The boil-off is compressed to 3.04 bars and intercooled in a separator by mixing with cold pressurized liquid ammonia. The liquid ammonia is separated and send to the tank, while the vapor are compressed further in the second compressor to 12.67 bars. Ammonia is cooled in a condenser and send to the separator as shown in figure 8.

Using DWSIM flowsheet, equipment size and material stream is calculated. This aids in calculating capital costs and annual manufacturing costs to install and operate ammonia storage system. The capital costs and manufacturing costs are shown in table 3 and table 4 respectively.

Note: The chill water system has been ignored in the calculations to simplify the approach.

Table 3: Cost of installing ammonia storage system

Equipment	Capital Cost (mil. \$)
Heat exchanger	0.16
Compressors	0.09
Tank (Inner and Outer)	5.33
Separator	0.07
C <sub>TM</sub>	6.65
C <sub>GR</sub>	8.39

### 3.3 CRYOGENIC AIR DISTILLATION

Cryogenic air distillation separates the air into different gases by exploiting the difference in liquefaction temperatures of nitrogen (77.4 K), oxygen (90.2 K) and argon (87.3K) [37]. Although the process is complex and with numerous fluid flow

Table 4: Operating Cost of different units

Process	HBP-HR	EHBP-HR	Ammonia Storage System	ASU	SMR-HR
Water Consumption (m3/yr)	13710 18	13710 18	6669	1851 57	5271 28
Cost of Water (mil. \$/yr)	3.06	3.06	0.01	0.41	1.18
Electric Consumption (MWh/yr)	71655	95684	185	2582 9	2371
Cost of Electricity (mil. \$/yr)	15.05	-	0.04*	5.42*	0.5
Natural Gas Consumption (tons/yr)	1860	-	-	-	7749 9
Cost of Natural Gas (mil. \$/yr)	0.82	-	-	-	34.04
Mass of Carbon dioxide (tons/yr)	4697	-	-	-	8660 4
Carbon Tax (mil. \$/yr)	0.05	-	-	-	0.97
Total Number of Operators	16	16	12	14	15
Labour Cost (mil. \$/yr)	0.87	0.87	0.65	0.76	0.81
Salvage Value (mil. \$)	5.58	5.21	0.84	2.45	7.61
Depreciation (mil. \$/yr)	2.51	2.34	0.38	1.10	3.42
Miscellaneous Costs (mil. \$/yr)	9.20	8.66	2.02	4.50	12.09
Manufacturing Costs (mil. \$/yr)	31.55	14.93	3.10*	12.20 *	53.01
			3.06	6.77	

\*represents systems supplied by grid energy.

, this process can supply high purity nitrogen, oxygen and argon with high product output [25] [38] [37]. Although the principle of cryogenic air separation is the same, the process has a lot of variation [37]. The process typically uses two or three distillation columns separate the gases in the air [19]. The second and third column are required for argon purification and will incur extra costs [19].

The plant has of a cold section and a warm section [37]. The warm section consists of compression, pre-cooling, drying and purification processes; while the cold section

consists of heat exchanger and distillation columns [37]. All equipment in cold section or cryogenic equipment are housed inside a "coldbox". This is to reduce cold losses by having container like steel structure filled with insulating material [37].

The ambient air is first fed to a mechanical filter to remove dust particles and then, compressed to about 6 bars using an intercooled staged compressor [39] [37] [40]. The compressed air is pre-cooled with chilled water (between 5°C -20°C) in a direct contact cooler [39] [37] [40]. This results in reduction of the moisture content of the compressed air. The chilled water is supplied by an evaporative cooler, where it exchanges heat with residual nitrogen-rich gas from the separation [39] [37] [40]. The cold and compressed air is finally fed into molecular sieve absorber, that is periodically loaded/regenerated, of warm section [39] [37] [40]. This results in removal of remaining water content, carbon dioxide and dilute amounts of numerous hydrocarbons before feeding into cold section of cryogenic air distillation [37] [40].

To simplify the approach the air is assumed prior to compression is to be clean. Cryogenic air distillation is simulated in DWSIM based on [41]. The simulation takes air at ambient pressure and at a temperature of 298.15 K. The mole fraction of incoming air constitutes of 0.78 of nitrogen, 0.21 of oxygen and 0.01 of argon. The compressed cold air is cooled to approximately liquefaction temperature in heat exchangers. The heat is exchanged from gas and liquid streams from the rectification process to incoming air. A fraction of the air is expanded in the turbine/throttling system to account for any thermal irreversibility in the process. The air is subsequently fed to high-pressure column (pressure is between 5.35 - 5.49 bars) where nitrogen collects at the top while oxygen at the bottom due to difference in the volatilities of the components. Nitrogen, from high-pressure column, is condensed and cooled by liquid oxygen sump from low pressure column in a combined condenser-evaporator. A fraction of the liquid nitrogen is used as a reflux for high pressure column and the rest is expanded and fed as reflux onto the top of low-pressure column. Moreover, the gases are separated further in low pressure column with typical pressure between 1.22-1.4 bars [37] [40].

Cryogenic air distillation varies from industry to industry as per requirement of that specific plant. However, some common equipment include multi-staged intercooled compressors, followed by cooling in heat exchangers, expansion in throttling valves and, separation in high pressure (HP) and low pressure (LP) distillation columns. The designs of cryogenic air distillation of industrial scale are propriety and are extensive making it beyond the scope of this paper, so a preliminary analysis will be based on process simulations based on available literature [39] [37] [40] [41].

The high purity nitrogen (99%) is produced from cryogenic air distillation at a rate 246.7 tons per day. Since HBP requires high nitrogen purity, this paper will not cover the argon distillation. Once the plant is simulated using flowsheet results, the equipment are sized and economic estimation is carried out. The capital costs and annual manufacturing cost to install and operate cryogenic air distillation unit are shown in table 5 and table 4 respectively.

### 3.2 STEAM METHANE REFORMING (SMR)

SMR process is utilized to produce hydrogen from methane and steam. The process is highly developed and mature technology in industrial sector for hydrogen

production. It produces approximately 48% of the world hydrogen production [42] and 72% of hydrogen for global ammonia production [5].

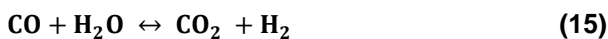
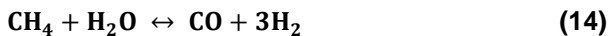
Table 5: Cost of installing cryogenic air distillation

Equipment	Capital Cost (mil. \$)
Heat exchangers	2.63
Compressors	8.21
Columns	6.04
Pump	0.02
C <sub>TM</sub>	19.93
C <sub>GR</sub>	24.51

SMR plant, using DWSIM, processes natural gas as discussed in [42] [43]. The natural gas is compressed to a pressure of 3 bars and passed through desulphurization unit to remove hydrogen sulphide and other acidic components. Then it is mixed with steam in a molar ratio of 1:2.4

The conversion of natural gas to hydrogen takes place in the reforming section. In the reforming section. In the reforming section gases are heated to 649°C since the process is highly endothermic as shown in equation 14. The production of hydrogen takes place in two sets of reaction as shown in equation 14 and 15. The heat for reforming reaction can be provided by external or internal means by burning natural gas or some gaseous fuel. The gaseous exit the reformer at 1000°C and results in the conversion of 97% of methane containing 17% of carbon monoxide.

This remaining carbon monoxide is further converted into hydrogen and carbon dioxide as shown in equation 15. The reaction is exothermic and thermodynamically favoured at low temperature. The water gas shift (WGS) is taken place in two reactors in series; high temperature shift (HTS) and low temperature shift (LTS) [42] [43].



The synthesis gas is cooled down to around 350°C and fed to HTS reactor. Moreover, the feed gases are further cooled to around 204°C and fed to LTS. In LTS reactor, residual CO conversion takes place and synthesis gas leaves the reactor containing 0.01% of CO.

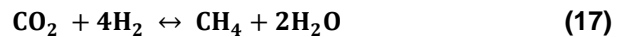
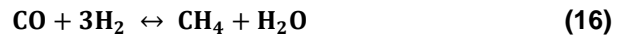
The process gas is cooled to condense the excess steam and subsequently, removing condensate from the process gas. Hydrogen purification can be achieved through CO<sub>2</sub> scrubber – methanation [42] [43]. The conventional scrubber absorbs carbon dioxide by using a chemical solvent (amine solutions) or a physical solvent (glycol dimethylethers).

The carbon dioxide scrubbers are complex and need to be represented on a model of its own. Since scope of this work focuses on SMR, the simulation, removing approximately 95% of carbon dioxide, considers carbon dioxide scrubber as a ‘black box’. The standard model for carbon dioxide scrubber is considered for cost estimation purposes by Even Birkelund [44].

The scrubber uses monoethanolamine (MEA) as an amine solution to absorb carbon dioxide. The technical and economic

details of the standard model are used from [44]. The capital cost of the scrubber includes the cost of pumps, heat exchangers, absorber column, and stripper column. The standard model in [44] removes 38 kg CO<sub>2</sub>/s is scaled down to 3.49 kgCO<sub>2</sub>/s required by SMR plant.

Methanation is next step after CO<sub>2</sub> scrubber, where trace amounts of carbon dioxide and carbon monoxide are converted to methane as shown in reaction below. This is to avoid any catalytic poisoning in ammonia synthesis. The reaction, as shown in equation 16 and 17, takes place around 300°C resulting in hydrogen purity of around 94% with small amounts of methane and steam. Methane is an inert gas for ammonia synthesis loop but moisture is removed before the gases are feed to ammonia synthesis loop [42] [43]. This gives hydrogen with a purity of 97.3%.



From the process discussed, the main process equipment include compressor, desulphurization unit, shift reactors, condensate removal unit, carbon dioxide removal or conversion unit and heat exchanger. The designs of industrial scale SMR unit are extensive making it beyond the scope of this thesis, so a preliminary analysis will be based on process simulations based on available literature [42] [43] [45] [46].

The high purity hydrogen required for HBP is produced from SMR plant at a rate of 53.3 tons per day. The process is simulated for required hydrogen flowrate and purity. Once the plant is sized and simulated, the stream data and flowsheet aids in sizing the equipment, and subsequently, economic estimation is carried out as discussed earlier. The capital costs and annual manufacturing cost to install and operate SMR unit are shown in table 6 and table 4 respectively.

Table 6: Capital cost of installing SMR-HR unit

Equipment	Capital Cost (mil. \$)
Heat exchangers	1.69
Compressors	0.52
Reactors	0.73
Separators	0.14
Boiler	20.0
Pump	0.04
CO <sub>2</sub> Scrubber	22.57
C <sub>TM</sub>	53.91
C <sub>GR</sub>	76.05

### 3.3 WATER ELECTROLYSERS

Water can be split into hydrogen at cathode and oxygen at anode electrochemically using electricity. This is an alternate route to produce high purity hydrogen other than the conventional SMR to produce hydrogen from natural gas for ammonia production. Currently, there are three types of electrolyser available; Alkaline electrolyser, solid oxide electrolyser and PEM electrolyser. Solid oxide electrolyser are not discussed in this work because it is currently under development and have no commercial availability in the market, while both alkaline and PEM electrolyser are commercially available [47] [48].

The alkaline electrolyser system is the most mature

electrolytic technology for the production of hydrogen. The alkaline electrolyser systems of today are able to produce 485 Nm<sup>3</sup>/h of hydrogen with an energy consumption of 4.85 kWh/Nm<sup>3</sup> of H<sub>2</sub> [49] [50]. The baseline case selected for the thesis is Norsk Hydro Atmospheric Type No. 5040 (5150 Amp DC), as the techno-economic details conforms to the current alkaline electrolyser are taken from [25] [50] [51].

PEM electrolyser system are a commercial technology to produce hydrogen as an alternate to alkaline electrolyser. PEM electrolyser stack of today is able to produce 10 Nm<sup>3</sup>/h of hydrogen with an energy consumption of 6.3 kWh/Nm<sup>3</sup> of H<sub>2</sub> [49] [50]. A baseline case is selected as PEM stack, Proton Hogen 380, for this work and the techno-economic details conforms to the current PEM electrolyser are taken from [50] [52].

The hydrogen required for 300 ton/day ammonia is around 53.3 ton/day, which corresponds to approximately 24887.54 Nm<sup>3</sup>/hr or 2220.83 kg/hr. This corresponds to approximately 51 modules of Norsk electrolyser, as shown in table 8, with a water requirement of 24735 kg/hr, or corresponds to approximately 2468 modules of PEM electrolyser, as shown in table 8, with a water requirement of 20723 kg/hr.

The capital cost, C<sub>GR</sub>, of Norsk electrolyser system can be calculated by incorporating 17% of direct installed capital to account for site preparation, engineering and design, project contingency and permit fees [53]. While, the capital cost, C<sub>GR</sub>, of PEM electrolyser system can be calculated by taking an installation factor of 1.12 and by incorporating 17% of direct installed capital to account for site preparation, engineering and design, project contingency and permit fees [52]. The capital costs and manufacturing costs, calculated as discussed earlier, for Norsk electrolyser and PEM electrolyser systems are show in table 7.

Moreover, the water must be purified and desalinated before feeding to alkaline electrolyser or PEM electrolyser. The high-water purity or total dissolved solids (TDS) requirements for alkaline electrolyser is less than 10ppm and for PEM electrolyser, it is less than 0.5 ppm [25] [54]. While the electrolyser units have a water purification section (reverse osmosis modules), they are not designed for desalination. Since the scope of this paper deals with coastal areas of Germany having access to sea and ocean water with a TDS range of 10,000-45,000 ppm, mechanical vapor compression (MVC) is selected to desalinate the water for electrolyser units [25] [19]. This is can be attributed to the MVC is all-electric plant with moderate investment having commercial viability for small and medium-scale system [55].

The MVC system are complex and need to be represented on a model of its own. Since scope of thesis focuses on the electrolyser system, the model put forward by Morgan et. al [25] is used. The model has a specific power consumption of 22.45 kWh/m<sup>3</sup> of water desalinated. Moreover, the grass root cost, C<sub>GR</sub>, can be calculated using equation 18 [25] [20].

$$C_{GR} = \left( \frac{3}{200} \dot{m}_{\text{water}} + 4.5 \right) \times 10^6 \quad (18)$$

Where,  $\dot{m}_{\text{water}}$  represents the distil water flowrate in ton/day. The capital costs and manufacturing costs, calculated as discussed earlier, for MVC systems are show in table 7.

Note: The selling price of oxygen is taken approximately \$150/ton [56] and the number of operators required for MVC, Alkaline electrolyser and PEM electrolyser system are based on Morgan et. al [25] and Ramsden et. al [51] [52].

Table 7: Capital and manufacturing costs of commercial water electrolyser systems

Equipment	Alkaline Electrolyser System		PEM Electrolyser System	
	MVC	Norsk Modules	MVC	PEM Modules
C <sub>GR</sub> (mil. \$)	13.40	136.03	11.96	232.85
Electricity Consumption (MWh/yr)	4442	955682	3723	1254332
Total Number of Operators	13	10	13	10
Labour Cost (mil. \$/yr)	0.70	0.54	0.70	0.54
Oxygen Production (tons/yr)	-	141376	-	140182
Oxygen Revenue (mil. \$/yr)	-	21.21	-	21.03
Salvage Value (mil. \$)	1.34	13.90	1.20	23.29
Depreciation (mil. \$/yr)	0.6	12.51	0.54	29.94
Miscellaneous Cost (mil. \$/yr)	2.81	20.96	2.60	34.65
Manufacturing Cost (mil. \$/yr)	4.12	12.81	3.85	44.11

### 3.4 ONSHORE WIND

Onshore Wind Power is a commercial technology to produce renewable energy as discussed in section 2.4. Recent studies by Fraunhofer Institute [57] show that the technology have a LCOE of €0.0399/kWh - €0.0823/kWh in Germany. The analysis was based on wind turbines with a rated capacity of 2 – 4 MW and 1800 – 3200 full load hours (FLH). The techno-economic details of onshore wind in Germany are based on [57].

The electricity generated using the wind far is used to drive air separation units, electrolyser stacks and HBP loop. LCOE is calculated using equation 19.

$$LCOE = \frac{I_0 + \sum_{t=1}^n \frac{A_t}{(1+i)^t}}{M} \quad (19)$$

Where, I<sub>0</sub> is the capital investment, A<sub>t</sub> are the operating costs, n is the project life (20 years), i is the discount factor taken as 3% [58] and M is overall energy production from the farm over the entire life of the project.

To simplify the approach, the tecno-economic data as per studies conducted by Fraunhofer Institute [57] is used. The coastal areas of Germany have wind speed is around 7.3 m/s at a hub height of 120m. The FLH are 3200 hours and the

capital expenditures are €1900/kWh. Table 8 shows the capital and operating expenditures of wind farms to supply different ammonia plants.

The electric power produced from wind farms is used to power electrolyser system, ASU, MVC, ammonia storage system and EHBP.

Table 8: Capital and operating costs of wind farms

Equipment	Wind Farm 2	Wind Farm 4
Power Production (MWh/yr)	1081822	1379753
Wind Farm Rated Power (kW)	338069	431173
LCOE (€/kW)	0.0404	0.0404
Capital Costs (mil. \$)	719.41	917.54
Operating Costs (mil. \$)	17.42	22.21
Process Units	Alkaline Electrolyser, MVC, ASU, EHBP-HR, Ammonia Storage System	PEM Electrolyser, MVC, ASU, EHBP-HR, Ammonia Storage System

### 3.5 LEVELISED COST OF AMMONIA

The levelized cost is an effective tool for comparing different production methods. LCOA helps in identifying different methods on the basis of investment required for producing a ton of ammonia. LCOA can be calculated using equation 20.

$$LCOA = \frac{I_0 + \sum_{t=1}^n \frac{A_t}{(1+i)^t}}{A_M} \quad (20)$$

Where,  $A_M$  represents the amount of ammonia produced in tons for the entire project life. The other expressions are same as that in equation 19. The LCOA for different ammonia synthesis systems, energy consumption per ton of ammonia and natural gas consumption per ton of ammonia are shown in table 9.

Table 9: LCOA of different ammonia plants

Plant	LCOA (\$/tNH <sub>3</sub> )	Process Units
Ammonia Plant 1	819	SMR-HR, ASU, HBP-HR, Ammonia Storage System
	714*	
Ammonia Plant 2	673	Alkaline Electrolyser, MVC, ASU, EHBP-HR, Ammonia Storage System
Ammonia Plant 3	635	PEM Electrolyser, MVC, ASU, EHBP-HR, Ammonia Storage System

\*represents LCOA without ammonia storage system, ASU and carbon tax.

### 3.6 FUTURE OF GREEN AMMONIA

The research and development of water electrolyser will aid in increasing the efficiency of the stacks, size of the stack and life of the stack. The techno-economic data of Alkaline and PEM electrolysers in 2030 as per a study conducted by E4tech [59] are used for predicting capital and operating costs of water electrolysers in 2030.

Similarly, the onshore wind farm power rating, capital cost and manufacturing costs in 2030 are based on studies conducted by Fraunhofer Institute [57]. The work by Fraunhofer Institute shows that the LCOE of onshore wind will be approximately €0.035/kWh in Germany by 2030 [57].

Moreover, there is a predicted increase by the year 2030 in the cost of carbon tax (€30/ metric ton of carbon), natural gas (€0.0322/kWh) and electric tariffs (€ 0.0422/kWh) [9] [60] [57]. This in turn can have a significant impact on the production of ammonia through SMR-HBP. LCOA of ammonia plants in 2030, can be recalculated using equation 20 and are shown in table 10.

Table 13: LCOA of different ammonia plants in 2030

Plant	LCOA (\$/tNH <sub>3</sub> )	Process Units
Ammonia Plant 1	798	SMR-HR, ASU, HBP-HR, Ammonia Storage System
	684*	
Ammonia Plant 2	917	Alkaline Electrolyser, MVC, ASU, EHBP-HR, Ammonia Storage System, Wind Farm 1
Ammonia Plant 3	1323	PEM Electrolyser, MVC, ASU, EHBP-HR, Ammonia Storage System, Wind Farm 3

\*represents LCOA without ammonia storage system, ASU and carbon tax.

## 4. CONCLUSION AND DISCUSSION

### 4.1 CONCLUSION

In this thesis, a preliminary cost estimation of 300 tonne per day ammonia plants is conducted that were built from “ground zero”. The thesis studies the technical and economic feasibility of producing ammonia powered by onshore wind and compare it with conventional ammonia synthesis process (using SMR and HBP).

- The results of work are summarized below.
- LCOA from SMR-HBP configurations are \$883/ton of ammonia and \$798/ton of ammonia.
- While LCOA using onshore wind farm to power alkaline electrolyser, ASU and EHBP-HR is \$917/ton of ammonia. Similarly, LCOA using onshore wind farm to power PEM electrolyser, ASU and EHBP-HR is \$1323/ton of ammonia.
- LCOA from alkaline and PEM electrolyser are higher due to higher capital costs of the electrolyser modules and, also due high electrical consumption by both electrolyser systems, higher capital costs of onshore wind farms.

The current economies show that SMR-HBP will continue to dominate ammonia synthesis market. This is can be

attributed to the fact that SMR-HBP have a lower energy requirement and less capital intensive when compared to green ammonia plants. Moreover, scale and production capacity of plants are big compared to the current feasibility studies, current demonstration plants or green ammonia plants in the past. However, the research and development in water electrolyser technologies can shift the market towards the green ammonia in the future [59] [17].

The research in alkaline electrolyser is directed towards improving the performance of catalytic/electrode materials [59] [61]. The concept of porous electrodes adhered to diaphragm helps in reducing the distance and energy consumption. This zero-gap configuration results in avoiding bubbles and non-conductive film over the electrodes (which increases the energy consumption and gas transport through the diaphragm) [59] [61]. These advanced concepts results in increasing the current density of the electrolyser by up to 0,5 A/cm<sup>2</sup> and pressures up to 30 bar [61].

Similarly, PEM electrolyser systems need to be built from inexpensive materials having longer life and efficiencies comparable to that of alkaline electrolyser [59]. The existing challenge is directed towards the replacement of platinum group metals with non-noble electrocatalysts and the replacement of noble catalyst in oxygen evolution reaction [59] [61]. Moreover, both water electrolyser systems need to be scaled up so they can gain economies of scale and be comparable to commercial SMR plants.

Furthermore, this transition from brown to green ammonia will also be depended upon a significant increase in carbon tax from \$11.21/ ton CO<sub>2</sub> to \$30/ton CO<sub>2</sub> (or more). However, the success of carbon capture and storage (CCS) both technically and economically can extend the time-period of this transition. The research is directed towards improving the performance of sorbents and chemicals, and removing the factors impeding the scaling up of CCS plant from lab scale to pilot scale. However, the barrier of safety plays a pivotal role in implementation CCS on a large scale [62].

Considering the reduction in capital costs (less than \$500/kW) and energy consumption of electrolyser (less than 4.5 kWh/kg of H<sub>2</sub>), followed by a significant increase in carbon tax in the future (more than \$50/ton of CO<sub>2</sub>), green ammonia can be more economically feasible in the future as shown by results below;

- LCOA in 2030 from SMR-HBP configurations are \$916/ton of ammonia and \$819/ton of ammonia.
- LCOA in 2030 using onshore wind to power alkaline electrolyser, ASU and EHBP, are \$709/ton of ammonia and \$673/ton of ammonia with heat recovery unit.
- LCOA in 2030 using onshore wind to power PEM electrolyser, ASU and EHBP, are \$670/ton of ammonia and \$635/ton of ammonia with heat recovery unit.

#### 4.2 RECOMMENDATION

This thesis is can be improved by;

- The ammonia reactor, reformer reactor, shift reactors and methanator reactor should be based on chemical kinetics in presence of catalyst similar to that based on commercial and industrial scale.
- ASU, SMR and HBP systems are quite complex with different configuration of the equipment and different operating conditions. The thesis approximates the process based on data provided by the literature and only incorporates the major equipment in the process. The auxiliary equipment costs are taken as a fraction of the

major equipment cost to simplify the cost estimation. However, a detailed economic and technical model should be developed as per industrial standards.

- All processes and systems are considered individually and in a steady state. Since ammonia synthesis process is complex, the system should be studied dynamically as it depended upon other process to produce hydrogen and nitrogen. A holistic approach should be used to model the system instead of an individualistic one. This can aid in heat and power integration to improve the system efficiency.
- Similarly, there are no connections to the grid for onshore wind farm powering the green ammonia plants. In reality, power is given to and taken from the grid for constant production of ammonia. Alternate scenario would be production of ammonia when power is produced from onshore wind farm only. In that case, ramp up and ramp down of processes and system should be considered.
- The models have not been validated, however, recently the successful operation of a demonstration plant in Oxford show that wind energy can be used to power electrolyser stacks, ASU, and HBP to produce ammonia.
- The scale of chemical plant is small compared to existing ammonia plants that operate at a production capacity of 3000 tonnes/day. To keep the comparison fair and in balance, one approach is to calculate LCOA of SMR-HBP plant producing 3000 tonne/day ammonia and compare it with LCOA of 300 tonne/day ammonia plant operating on onshore wind and water electrolyser.
- Another factor to consider while doing LCOA calculation in the future is to include the projected economies of CCS technology. This approach can portray a clear picture whether green ammonia or brown ammonia will dominate the market.

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