

3.0

Large scale hydrogen production and storage technologies

This section sets out a robust comparison of the current hydrogen production technologies which are commonly referred to as 'large' scale. These include:

- Water electrolysis
- Natural gas reforming
- Coal gasification
- Ammonia production and cracking

In addition, potential inter-seasonal hydrogen storage technologies have been evaluated, with the focus being on salt caverns and ammonia as the most technically developed existing storage solutions.

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3.1 Overview

There are several production routes for hydrogen, as illustrated in [Figure 3.1](#). The choice depends on volume of production, cost, type of available feedstocks and utilities, and the cost. In a true large-scale hydrogen economy, it is likely all large-scale production methods (and potentially many more) will play a part.

However, for the purposes of this study, the most viable supply methods have been analysed in the context of their applicability to the H21 NoE demand parameters (see [Section 2.0](#)) and timescales.

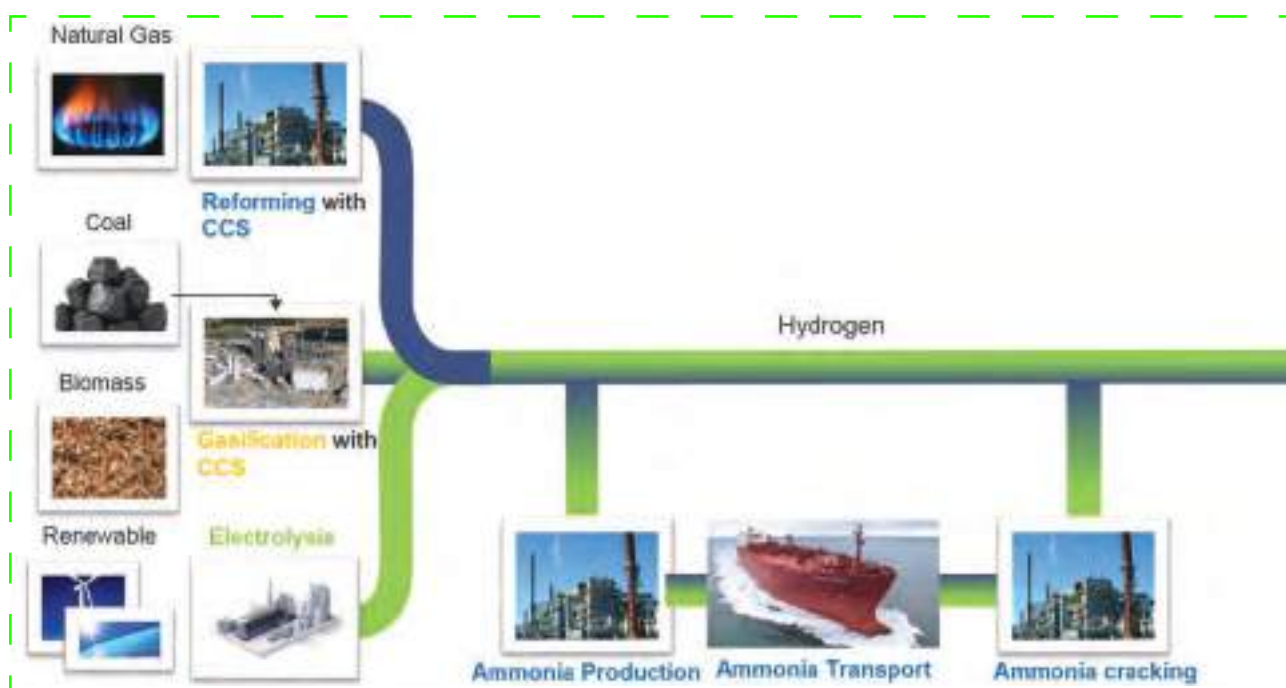
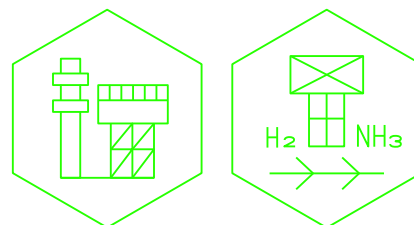


Figure 3.1: Potential production routes to hydrogen

→ In [Section 2.0](#) it was estimated the peak annual demand to be 85.12TWh for H21 NoE, which indicates that the hydrogen production capacity should be in the range of 10-15 GW. With commissioning between 2026-35, 1 to 2 GW of hydrogen production capacity would have to be installed per year (see [Section 6.0](#)). Based on these parameters the task was to identify the most appropriate form of hydrogen production technology to meet the demand requirements, i.e. at 1-2 GW scale.

A hydrogen production solution should fulfil the following key requirements:

- Highest possible energy efficiency to minimise losses and cost of energy
- Lowest possible capital and operating cost
- Lowest possible CO₂ footprint to meet UK emission targets
- Technology and supply chain maturity that allow for commissioning by 2026
- Minimum requirements for use of water resources and land
- Excellent safety track record and reliability for robust security of supply

Equinor undertook a detailed analysis of the major hydrogen production technologies. The first step was to compare at equal basis, i.e. the selected scale in the range of 1.3-1.5 GW and generic/equal site conditions (Section 3.1-3.7). In the second step these technologies were further assessed for the H21 NoE supply chain considering all elements in the value chain that needed to be established (Section 3.8). The selected technology and concept is further detailed for H21 NoE in Sections 4.0 and 6.0.

There will be minor differences between the frame condition in which the technology have been assessed compared to the final H21 design, e.g. in the generic technology comparison and H21 system consideration it is assumed to import power needed for the ATR based Hydrogen Production Facility whereas in the final H21 design the power is produced internally at Hydrogen Production Facility. This results in a slightly different CO₂ footprint and efficiency.

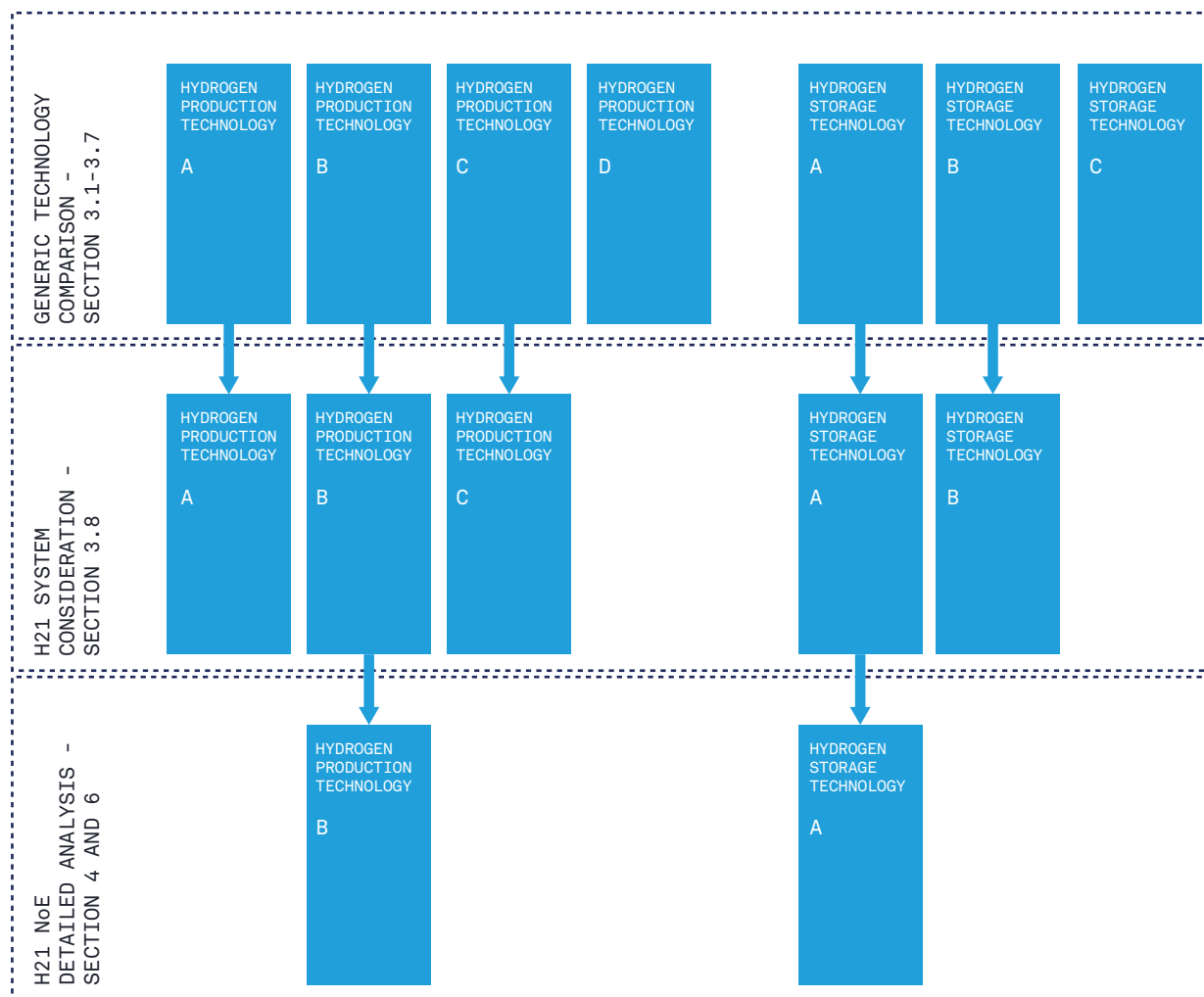


Figure 3.2: H21 NoE technology and concept assessment process

A summary of comparisons across technologies is presented at the end of this section with explanation for the method selected for the NoE area of conversion.

3.2

History – global H₂ production

Hydrogen is the lightest molecule and the first atom in the periodic system. The development of the universe starts with hydrogen and from there all other atoms originate.

Hydrogen is the most abundant molecule in the universe and is the fuel that drives the sun. However, on earth free hydrogen does not exist in significant amounts and needs to be produced from other hydrogen containing sources such as water (H₂O), natural gas (CH₄), hydrocarbons (CnH_{2n+2}), etc.

Methods for producing hydrogen in the form of towns gas have been known for more than 200 years. Towns gas was produced by heating coal to produce gas containing primarily hydrogen, methane and carbon monoxide. In the late 1800s industrial hydrogen production from water electrolysis started to develop.

Industrial large-scale hydrogen production started after World War I because of the wide scale application of the Haber-Bosch process (patented by IG Farben) to produce ammonia for fertiliser manufacture. The Haber-Bosch process was based on fixation of hydrogen with nitrogen into ammonia. The Norwegian company Norsk Hydro collaborated with IG Farben in the 1920s on large ammonia plants based on hydrogen from water electrolysis.



Figure 3.3: Water electrolysis plant 135MW in Rjukan, Norway (1948)

IG Farben also patented a nickel catalyst for steam methane reforming purposes. During the 1920s both IG Farben and the UK company ICI (later Johnson Matthey Process Technologies) worked on development of a steam methane reformer. In 1930 IG Farben, Standard Oil and ICI entered into an agreement to share development. The first steam methane reformers were started-up in the early 1930s in Bayway, USA by Standard Oil and in 1936 the first steam reformer in Europe was put into operation at the Billingham site by ICI. Over the 30-year period between 1930 and 1960 steam methane reforming went through rapid development. This included the introduction of the autothermal reformer (ATR), operating at higher pressure and catalyst development resulting in improved energy efficiency.

Today 95% of all global hydrogen production is based on reforming natural gas and oil (naphtha) or gasification of coal, see [Figure 3.4](#).

Current global hydrogen production is 60 Mtpa (2,400TWh). The demand per industry segment is shown in [Figure 3.5](#). The biggest proportion (65-70%) is used directly for synthesis of new products, e.g. ammonia and methanol. Another significant proportion goes to refineries for desulphurisation of road transport fuels to meet low sulphur regulations and synthetic fuels.



Global production
60 Mt

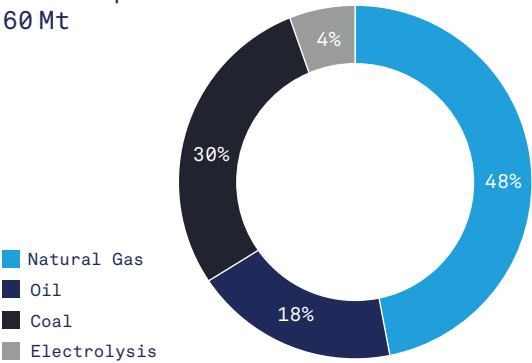


Figure 3.4: Global hydrogen production based on feedstock

Global hydrogen
demand 60 Mt

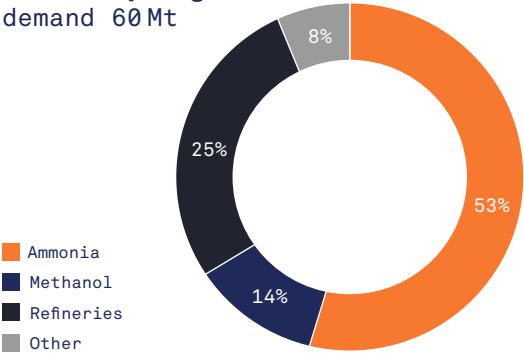


Figure 3.5: Global hydrogen demand

It should be noted that the majority of hydrogen production capacity has been installed over the last 40-50 years. To appreciate the scale of production the following statistics should be considered

- 60 Mt of hydrogen are equal to 280 GW hydrogen output (HHV). This equates to 280 times Leeds or 5.5-7 GW installed annual capacity per year assuming an average of 40-year lifetime of which 65% or 3.5-4.5 GW per year have been ATR/SMR technologies
- 40 Mt of hydrogen produced via SMR/ATR equates to the capacity of 500 SMRs and 300 ATRs/secondary reformers globally
- The current hydrogen industry is a part of the oil & gas and power industries. It shares the same supply chains. Therefore, it can be reasonably assumed, that the whole industry has the logistical strength to develop new hydrogen capacity at 5-10 times the current pace, i.e. 30-70 GW per year.

Furthermore, between 1965 and 1985 ammonia production increased from 20 Mtpa to 100 Mtpa. The growth of 80 Mt (500 TWh) in 20 years equals an installed capacity of 25 TWh every year of 3 GW on hydrogen production (see Figure 3.6). As a point of reference, this historic growth in hydrogen production for the ammonia sector is three times the annual growth in hydrogen capacity required by the H21 NoE concept for the period 2026-35.

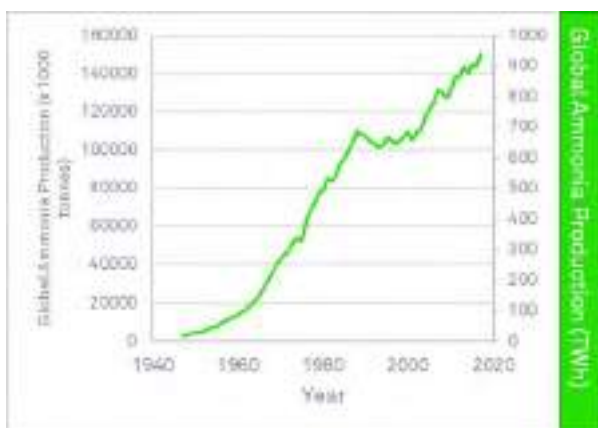


Figure 3.6: Global ammonia production from 1946 to 2007



3.3

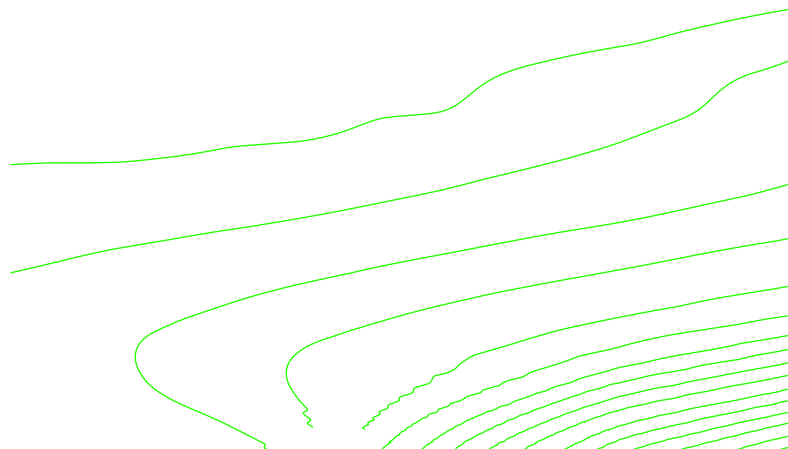
Water electrolysis



Water electrolysis remains the most efficient route to hydrogen from renewable electricity sources and is therefore expected to play an increasing role as energy systems become more dependent on renewable sources like wind and solar.

This is already emerging in the transport sector and for grid balancing purposes, or where smaller, distributed electrolyser units can be installed.

Water electrolysis is therefore a potential solution for decarbonisation of the energy segment and Equinor analysed the merits of the technology based on open literature combined with their own in-house experience from being the owner of the world leading electrolyser company NEL until 2011/12.



3.3.1 Water electrolysis global production and locations

Industrial production of hydrogen by water electrolysis started at the end of 19th century. From the beginning, the energy supply necessary for electrolysis was based on renewable energy in the form of hydroelectric power.

The information on capacity, year of installation and decommissioning differs between publications, but it is generally agreed that the Rjukan plant was the first with a capacity of 27,900 Nm³/h or 2,380 MWh/d. In Norway water electrolysis established an important basis for industrialisation and economic growth due to the cheap and abundant hydroelectric power.

In addition, Norway has practically unlimited amounts of freshwater.

Whilst the Rjukan plant was one of the largest electrolyser facilities ever built, it is important to contextualise on a heat demand scale. This volume of hydrogen would only account for 1% of the H21 NoE average demand.

An overview of large-scale plants worldwide is shown in [Table 3.1](#). Some of the largest plants currently in operation are in Aswan, Egypt and in Nangal, India.

LOCATION	CAPACITY (NM ³ /H)	CAPACITY (MW)	POWER INPUT (MW)	EFFICIENCY (%)	MANUFACTURER	STATUS
Zimbabwe/Kwe-Kwe	21,000	74.6	100	75	Lurgi	Installed 1975
Norway/Rjukan	27,900	99.0	142	70	Norsk Hydro	Installed 1927-65 Decommissioned
Norway/Glomfjord	27,100	96.5	142	68	Norsk Hydro	Installed 1953. Decommissioned
Egypt/Aswan	32,400	115	160	72	BBC/DEMAG	Installed 1960-77
Peru/Cusco	4,700	15.6	22	71	Lurgi	Installed 1965
Canada/Trail	21,000	74.5	No info	No info	Cominco	Installed 1939
India/Nangal	30,000	106.5	142	75	De Nora	Installed 1958

Table 3.1: Large scale water electrolysis plants worldwide. Source: Smolinka (2012)

From the 1960s and onwards, water electrolysis was gradually out-performed by methane reforming. In Norway, the Rjukan and Glomfjord plants were decommissioned several decades ago. Today less than 4% of the global production of hydrogen comes from water electrolysis. The industrial use of water electrolysis is of smaller scale, typically less than 70 MWh/d for use in processes such as supplying hydrogen for fat hardening in the food industry.

In the EU the Fuel Cells and Hydrogen Joint Undertaking (FCHJU) was established in 2008 and is an industry-led public-private partnership aiming to bring clean, efficient and competitive fuel cell and hydrogen solutions to the market. An FCHJU study from 2014¹ concluded that electrolysis are expected to become more widespread in energy applications. Hundreds of MW of installed capacity is predicted across Europe between 2020 and 2025 if costs are reduced and system performance improved.

From 2008 to 2014 the FCHJU have spent circa €1bn to support water electrolysis and fuel projects and is expecting to spend another €1.33 bn from 2014 to 2020. Around 50% is allocated to so-called PEM (Proton Exchange Membrane) technology and 12% to Alkaline Electrolyser technology (AEL) the remainder being fuel cells. The increased interest from commercial players is resulting in a growing number of demonstration projects, especially across Europe, attempting to improve technology in size and performance.

¹ Bertuccioli, L. et al., "Development of Water Electrolysis in the European Union", FCHJI, 2014.

3.3.2 Description of the technology

Reference is made to the literature for detailed reviews of water electrolysis, such as Buttler and Spliethoff², and Smolinka. Ursua³ as well as Zeng and Zhang⁴ offer analyses of AEL systems and Carmo⁵ a detailed review of PEM technology.

Water electrolysis is the process of splitting the water molecule into hydrogen and oxygen by use of electricity.



The overall energy requirement of the reaction is around 285 kJ/mol at typical operating conditions for AEL and PEM. The theoretical minimum voltage required for electrolysis to take place when all energy is supplied electrically is around 1.48 V. In practice the voltage applied must be higher to overcome various resistances, including ohmic resistances, the activation energies of the hydrogen and oxygen formation as well as transport-related resistances. According to Buttler and Spliethoff, the required cell voltage of commercial electrolyzers is minimum 1.7 V for AEL and 1.65 V for PEM.

The basic building block of the electrolytic hydrogen production system is the electrolysis cell, which consists of a cathode and anode separated by a membrane (PEM) or a diaphragm (AEL). Several cells can be connected in series to produce a larger electrolysis cell stack, as illustrated in Figure 3.7 and Figure 3.8. Manufacturers often use a basic module, which may contain one or several stacks in parallel, and then configure commercial electrolyzers with several of them to achieve the desired production rate.

In the case of AEL the stack is immersed in a liquid electrolyte, usually aqueous potassium hydroxide (KOH), whereas the membrane used in PEM is selectively hydrogen ion conducting and acting as the electrolyte.

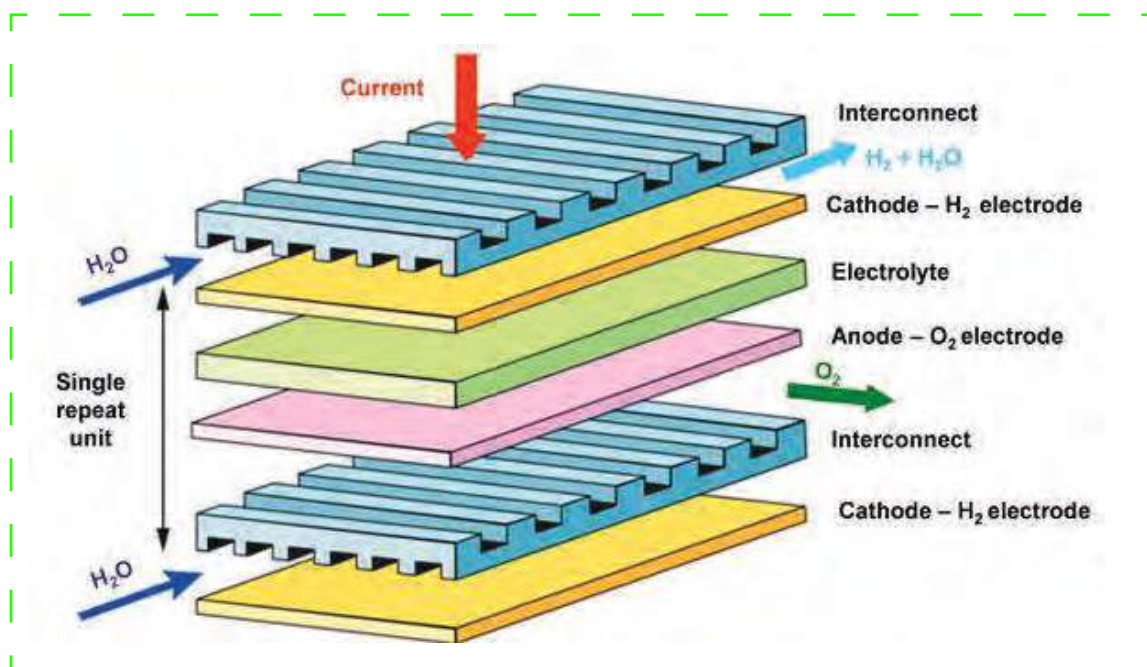


Figure 3.7: PEM electrolysis cell unit in cell stack (Source: Agral EU project)

2 A. Buttler, H. Spliethoff, "Current status of water electrolysis for energy storage, grid balancing and sector coupling via power-to-gas and power-to-liquids: A review". Renewable and Sustainable Energy Reviews 82 (2018) 2440-54.

3 Ursua et al "Integration of commercial alkaline water electrolyzers with renewable energies: Limitations and improvements", Int J Hydrogen Energy 41 (2016) 12,852-12,861

4 K. Zeng, D. Zhang, "Recent progress in alkaline water electrolysis for hydrogen production and applications", 2010.

5 M. Carmo et al., "A comprehensive review on PEM water electrolysis", Int. J. Hydrogen Energy 38 (2013) 4,901-4,934



Figure 3.8: PEM electrolysis cell stack (Source: Hydrogenics)

The hydrogen production system based on water electrolysis will also include additional equipment, and it varies between publications as to what is included. The recently published “EU harmonised terminology for low temperature water electrolysis...” aims to provide clear definitions and system boundaries⁶. An electrolysis system shall, in addition to the electrolysis stack, include:

- A power supply system, including distribution, rectifier and control system
- A water conditioning system, including water feed pump and a deionised water production unit
- In the case of PEM, an extended water conditioning system with make-up water storage tank, additional water purification unit, gas-water separators and demisters
- In the case of AEL, a liquid electrolyte supply system, including storage tank, pump and heat exchanger, gas-electrolyte separators and demisters
- Gas cooling
- Gas purification
- Process cooling
- Process utilities including water circulation pump
- Gas compression to required pressure

A rectifier is usually necessary to convert the AC voltage supply into DC current input. This conversion stage is outside of the system boundary. According to Ursua et al. the AC-DC and DC-DC power conversion stages needed between a windfarm and an electrolyser may have efficiencies as low as 70%.

Examples of AEL and PEM systems taken from Nel are shown in Figure 3.9 and Figure 3.10, respectively⁷. Note that these system descriptions are not in full agreement with the EU terminology. The water purification step is lacking, whereas the AC-DC converter is included.

⁶ Tsotridis, G. and Pilenga, A., “EU harmonized terminology for low temperature water electrolysis for energy storage applications”, 2018

⁷ http://nelhydrogen.com/assets/uploads/2017/01/Nel_Electrolyser_brochure.pdf

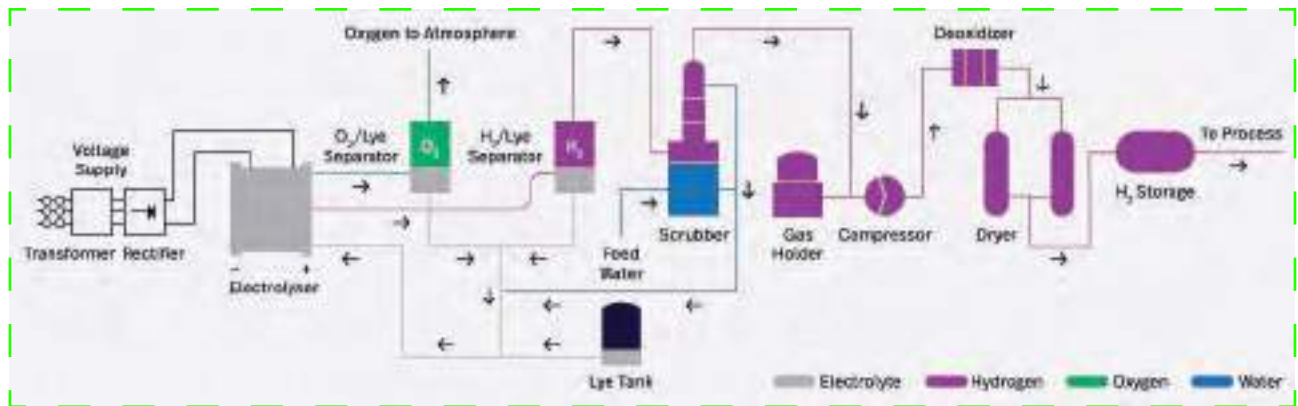


Figure 3.9: Alkaline water electrolysis production system. Source: Nel

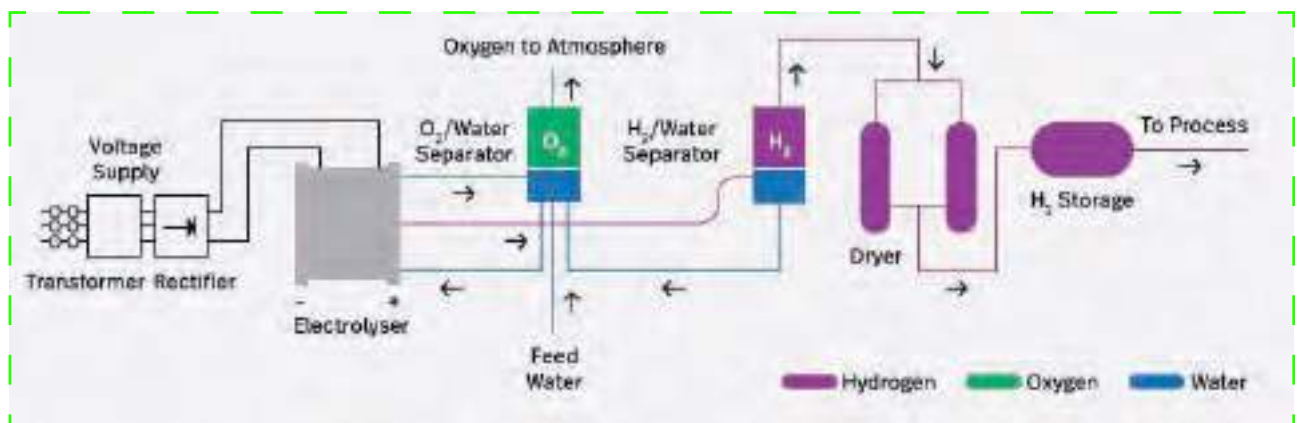


Figure 3.10: PEM water electrolysis production system. Source: Nel



3.3.3 Supply chain maturity

In 2014 the FCHJU described in the electrolyser industry as small and fragmented. This is still true, but the commercial interest for water electrolysis for energy applications is on the rise. H2-international, a journal on hydrogen and fuel cells, reported last year that the market for electrolysers is growing, both in terms of manufacturers and capacity ranges⁸. It identified more than 30 suppliers, although many of these are small with very limited supply chain capacity. Buttler and Spliethoff describes the market as very dynamic with several mergers and acquisitions in recent years. A not exhaustive overview of commercial players with geographic locations is shown in Figure 3.11.

There are several companies actively working towards some significant developments which could make step change contributions to the electrolyser market. These contributions coupled with growing renewable deployment (see Section 3.8), make the longer-term outlook and application for large scale electrolyser technology bright. However, this is likely post 2035 in the context of UK gas network conversion or other truly large-scale (GW) production.



Figure 3.11: Commercial players targeting electrolytic hydrogen for the energy sector

- a) NEL acquiring Proton Onsit
- b) HydrogenPro Tianjin strategic alliance 2018
- c) Hydrotechnik offers PERIC products in Europe
- d) Thyssenkrupp-McPhy strategic alliance 2015
- e) Thyssenkrupp and Asahi Kasei are major chloro-alkali industry suppliers

The most striking new trend is perhaps that leading chloro-alkali electrolyser manufacturers Thyssenkrupp⁹ and Asahi Kasei¹⁰ are aiming to take positions in the water electrolyser market for large scale energy applications. In contrast to the traditional water electrolyser manufacturers, who may have delivered up to a few hundred electrolyser units, Thyssenkrupp and Asahi Kasei have delivered more than 100,000 units – and therefore presumably have better manufacturing and delivery capabilities, which may enable faster deployment and lower investment costs.



Furthermore, whereas energy efficiency has been of minor importance for commercial applications of water electrolysis during the past decades, the chloro-alkali industry has continuously been competing against each other on capital cost and energy efficiency, building experience from plant support to thousands of installations.

In July 2018 Thyssenkrupp announced their large-scale water electrolysis technology, based on proven technology from the chloro-alkali business segment¹¹. They offer large systems based on 5 and 20 MW modules, worldwide EPC turnkey execution capabilities and an established 600 MW/year highly automated supply chain. Information is scarce, but the concept is likely to be already proven. A demonstration plant to produce green ammonia, integrating their proprietary ammonia technology with the water electrolyser, is planned in Port Lincoln, Australia¹².

Asahi Kasei is developing a 10 MW class unit. The target is to demonstrate a 10% more efficient system than the existing ones as well as fluctuating power adaptability, i.e. 1,000 hours of demonstration has already been done in Japan with partial scale-up to the large system¹³. In August 2017 they announced receiving an order from Toshiba Energy Systems to purchase a single unit of 10 MW to be installed in the Fukushima Hydrogen Research Field¹⁴.

In the PEM arena a handful of suppliers are working to demonstrate their technologies on the several MW scale including Siemens, Hydrogenics and ITM Power.

Despite these advances the supply chain must be considered immature when considering a multi-GW regional deployment of hydrogen production technology from 2026 onwards which is the basis for the H21 NoE concept with the success of such a system needing to be based on a guaranteed design and build capability starting in 2019 ready for commission in 2026.

9 Water electrolysis: Power to gas – thyssenkrupp AG

10 <https://ecs.confex.com/ecs/232/webprogram/Paper104839.html>

11 https://www.thyssenkrupp-industrial-solutions.com/en/press_detail_48130.html

12 <https://ammoniaindustry.com/thyssenkrupps-green-hydrogen-and-renewable-ammonia-value-chain/>

13 The tested stack contained 4 full size electrolysis cells. The large system producing 2000 Nm³/h will need 320 cells

14 <https://www.asahi-kasei.co.jp/asahi/en/news/2018/e180810.html>

3.3.4 Key performance

Supplier data on general performance and energy consumption in particular is not readily available. In addition, no commercial and validated tools/software exists to evaluate the electrolyser technology as is the case with reforming and gasification processes. The following summarises key information that could be found in brochures and other publicly available sources.

System efficiency

It is important to be aware that the efficiency of an electrolyser will depend on the hydrogen production rate. In practice there is a trade-off between efficiency and production rate. The higher hydrogen production, the higher the cell voltage or energy consumption per volume of hydrogen produced (i.e. the lower the efficiency). Manufacturers will typically optimise their product for a single or nominal production rate. Historically, this is how large-scale plants coupled to a stable hydroelectric power source were operated. The situation is different for more intermittent sources like solar and wind where the performance across a range of loads becomes important.

Targeting high efficiency for a given system will also tend to increase investment cost because a larger electrolyser is required. But the operating cost will decrease because of the lower energy consumption, less efficiency degradation and increased lifetime of the electrolyser stack.

It should be noted that direct comparison of performance figures of different products may be misleading as different suppliers highlight the performance of their products in different ways. For example, if a company wants to show a high H_2 production capacity, it could be that this production statistic is presented as the focus rather than the likely low energy efficiency and very high stack degradation rate.

Figure 3.12 gives an overview of efficiency figures for commercial, or close to commercial electrolysers that could be found in brochures or other open sources.

The selection of companies is a mix of well-established water electrolyser manufacturers like McPhy Nel and Hydrogenics, HydrogenPro (a Norway-based start-up with Chinese partner Tianjin) and new companies to the market like Siemens, Asahi Kasei and Thyssenkrupp. In addition, smaller companies like ITM Power and Giner ELX may have a competitive technology. Some efficiency figures were given on a stack level and others on a system level, but the system boundaries are not necessarily the same.





Figure 3.12: Energy efficiency of selected (pre) commercial electrolyzers. Efficiency is either given directly by the manufacturer or calculated as the ratio between the HHV (3.55 kWh/Nm^3) of hydrogen and the specific energy consumption multiplied by 100%.

As expected the system efficiencies are consistently lower than the stack efficiencies, averages being 69 and 79% respectively. It appears that AEL manufacturers often give stack efficiencies whereas PEM manufacturers typically use system efficiencies. There is no significant difference between AEL and PEM reflected in these.

Asahi Kasei and Thyssenkrupp claim the highest AEL stack efficiency (83%). Siemens reports the highest PEM system efficiency (75%) for their recently announced 3 MW SILYZER 300¹⁵. Siemens claim the SILYZER series of products are proven, but there are limited references. The SILYZER 200 (also shown in Figure 3.12) has been tested at Energiepark Mainz and some results have been published have been published. These show significantly lower energy efficiencies than claimed for the SILYZER 300¹⁶.

Usually the specific energy consumption is given on a stack basis. Some figures have been collected in Table 3.2 and Table 3.3 for AEL and PEM, respectively. 0.6 kWh/Nm^3 was added to estimated energy requirement at system level where such data were lacking based on the Buttler and Spliethoff estimated utility and rectifier energy requirement of $0.4\text{--}0.8 \text{ kWh/Nm}^3$.

The system energy consumption is $5\text{--}5.4 \text{ kWh/Nm}^3$. The four AEL systems have rather similar performance, only Hydrogenics providing data on the system level. The Nel system appears more efficient at stack level, but it delivers hydrogen at 1 bar, whereas the others deliver at around 30 bar. Efficiency around 70% (HHV), seems a plausible number for AEL.

Supplier data on PEM shows energy consumption figures that are quite similar, ranging from 5.3 to 5.6 kWh/Nm^3 , and hence somewhat higher than for AEL. The Siemens numbers appear reliable as system efficiency figures. The energy consumption vs hydrogen output from the Energiepark Mainz testing has been published (see Buttler and Spliethoff). The average system efficiency is 66% (HHV).

¹⁵ https://www.siemens.com/content/dam/webassetpool/mam/tag-siemens-com/smdb/corporate-core/sustainable_energy/hydrogensolutions/broschüren/ct-ree-18-047-db-silyzer-300-db-de-en-rz.pdf

¹⁶ https://www.utilities.nl/wp-content/uploads/2017/12/Siemens-FME-Industrie-en-Energie-2017_H2.pdf

COMPANY AND AEL PRODUCT	NOMINAL H ₂ RATE (NM ³ /H)	NOMINAL POWER (MW)	MAX OUTPUT PRESSURE (BAR)	ENERGY CONSUMPTION STACK (KWH/NM ³ H ₂)	ENERGY CONSUMPTION SYSTEM (KWH/NM ³ H ₂)	SYSTEM EFFICIENCY % (HHV)	LOAD FLEXIBILITY (%)	WATER CONSUMPTION (L/NM ³)
Nel A-485	300-485	2.2 (peak)	1 ^{a)}	3.8-4.4	4.4-5 ^{a)}	71-80	15-100	0.9 (feed water)
McPhy McLyzer 400-30	400	2	30	4.5	5.1 ^{b)}	70		
Hydrogenics Hystat-100-10	100	0.5	27		5-5.4	66-71	5-100	<1.7 (tap water)
HydrogenPro - Tianjin	800	3.5	30	4.4	5.1 ^{b)}	70		

Table 3.2: AEL performance data from suppliers

- a) Nel gives max pressure of 200 bar in their brochure, but that is with an external compressor installed. Energy consumption is only given at stack level. 0.6 kWh/Nm³ is added to give a system level estimation at the highest H₂ flow
- b) Only stack energy consumption given. 0.6 kWh/Nm³ is added to estimate energy consumption at system level

COMPANY AND PEM PRODUCT	H ₂ RATE (NM ³ /H)	RATED POWER (MW)	MAX OUTPUT PRESSURE (BAR)	ENERGY CONSUMPTION STACK (KWH/NM ³ H ₂)	ENERGY CONSUMPTION SYSTEM (KWH/NM ³ H ₂)	SYSTEM EFFICIENCY % (HHV)	LOAD FLEXIBILITY (%)	WATER CONSUMPTION (L/NM ³)
Hydrogenics HYLYZER 600	620	3	35	4.7	5.3 ^{a)}	67	No data found	<1.4 (tap water, HyLYZER®-400-30)
Giner ELX Allagash	30-400	2	40	5 ^{b)}	5.6	63	7.5-100	
Nel/Proton M400	417	2.2	30	4.8	5.4 ^{a)}	66	0-100	0.9 (feed water)
ITM Power HGas	630	3.5	80		5.5 ^{b)}	65	No data found	
Siemens Silyzer 200	225	1.25	35		5.56 ^{c)}	65-70	No data found	1.5 (tap water)

Table 3.3: PEM energy performance data from suppliers

- a) 0.6 kWh/Nm³ added to estimate system energy consumption
- b) Number taken from Buttler and Spliethoff
- c) Energy consumption calculated from rated power and nominal H₂ flow. Gives 64% efficiency

Water requirements

Little attention has been given to the water consumption of water electrolyzers, yet this may be an important issue in most parts of the world. Four companies provide open source data on tap water consumption including Nel (0.9 l/Nm³ H₂, feed water – may not be total), Hydrogenics (<1.4 l/Nm³ H₂, PEM), Siemens (1.5 l/Nm³ H₂) and Thyssenkrupp (<1 l/Nm³ H₂). The figures for AEL and PEM appear similar.

Hence, it takes roughly 1 litre of water to produce 1 Nm³ of hydrogen, compared to 0.4 litre for methane reforming. The average water volume required to supply hydrogen for H21 NoE based on water electrolysis would then be around 10 million litres per day. Desalination of sea water may be necessary. Process water or boiler feed water quality must be assumed, adding significant cost if seawater is used¹⁷.

Electrolysis plant footprint

There is one published example of a potential alkaline electrolyser plant from Nel based on their proven and referenced alkaline model 485 shown in Figure 3.13. It gives a footprint of 550 m x 350 m = 192,500 m² for a plant based producing 120,000 Nm³/h or 426 MW. This equals a specific footprint of 45.2 ha/GW.

Hydrogenics on the other hand estimates the footprint of a 100 MW plant based on PEM HYLYZER® 600 to 35 m x 50 m = 1,750 m². ITM estimates a similar facility to have a 3,500 m² footprint which is of the same order of magnitude. This may give the impression that PEM is much more compact than AEL. However, McPhy estimates that a 100 MW plant based on their alkaline McLyzer will need less than 4,500 m² which is comparable to the PEM estimates. The published numbers give a specific footprint between 2.5–7.0 ha/GW. It should however be noted that the exact scope and level of maturity of the projects is unknown. With a new technology, like the large-scale PEM electrolyser, which is still to be built, there is always considerable uncertainties around published figures.

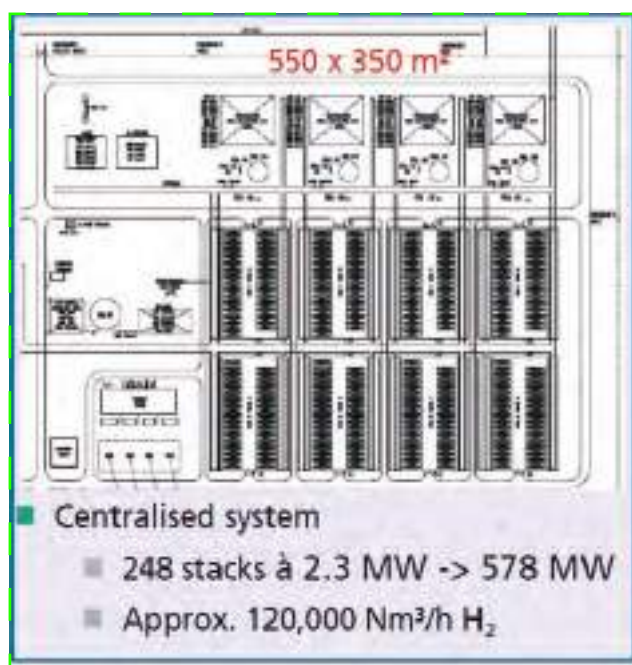


Figure 3.13: Suggested footprint of a 578MW_{e1} electrolyser plant from Nel¹⁸

¹⁷ <https://link.springer.com/article/10.1007/s40095-014-0104-6>

¹⁸ <http://www.fch.europa.eu/sites/default/files/2%20Water%20Electrolysis%20Status%20and%20Potential%20for%20Development.pdf>

Lifetime

According to Buttler and Spliethoff typical lifetimes of about 20 years for PEM and between 30 and 50 years for stationary operated AEL can be assumed. The lifetime of a stack depends on the acceptable efficiency reduction caused by degradation.

US Department of Energy reports a plant life of 20 years and electrolyser cell replacement intervals of 7 years for PEM¹⁹. Stack lifetimes in the Felgenhauer and Hamacher data ranges from 6 to 11 years for AEL and 8 to 11 years for PEM. No efficiency reduction criterion was set and there is no correlation between efficiency degradation and stack lifetime given by the different quotes. PEM stacks have higher degradation rates but are reported to have the same lifetime as AEL. In two cases the efficiency degradation is 2.5%/year for a stack lifetime of 11 years. This corresponds to an efficiency degradation of 29% at replacement. This might not be acceptable from an efficiency standpoint. If we assume that 10% efficiency degradation is acceptable (and that this is the determining factor for stack replacement) most AEL lifetimes fall in the range of 7 to 11 years and PEM in the range of 4 to 9 years.

To compensate for efficiency losses the power supply must be designed with additional capacity (e.g. 10-15%) with worst condition being a peak year at the end of stack lifetime.

Again, it is important to note that the Felgenhauer and Hamacher lifetime and probably all other lifetime figures are given for stationary operation at nominal current density. Many authors, including Smolinka, Bertuccioli and Buttler and Spliethoff point to the unknown impact of dynamic load operation on lifetime for both PEM and AEL. Ursua et al (2016) reports that degradation occurs when cell voltage drops below a minimum value for more than a few minutes. This strongly suggests that lifetime will be shorter than given above if the electrolyzers are coupled to a fluctuating energy source. There are ways to keep the cell voltage above the minimum value such as a battery providing a protective current, however, this comes with an energy and cost penalty. According to Smolinka, thermal cycling caused by repeated start-ups and shutdowns may also destroy seals and further reduce the lifetime.

Pressurised versus atmospheric operation

The potential advantage of pressurised electrolysis is a topic of controversy. The preferred choice in academic research and among the established water electrolyser manufacturers appears to be pressurised systems. Advantages are equipment compactness and the possibility of eliminating external compression stages. However, it is worth noting that the electrolyzers of the much larger established chloro-alkali industry are atmospheric.

The standard pressure of AEL and PEM is mostly below 30-50 bar, although higher pressures have been demonstrated. PEM has an advantage here, due to less cross permeation of hydrogen. But higher pressure leads to higher degradation rates triggering the need for more expensive materials or reducing lifetime. Pressurised operation will also require additional safety measures, both in design and operation. It will furthermore reduce the load range compared to atmospheric systems to avoid unacceptable cross permeation of hydrogen.

¹⁹ James, B.D. et al., "Hydrogen production pathways cost analysis 2013-2016", DOE-strategic-analysis 2361, 2016

Capital cost

In general, it is difficult to assess the credibility of published data of capital cost. Most published cost data originates from electrolyser suppliers and academia. So far, no real project costs have been reported from electrolyser plant owners and operators. Therefore, cost data, and in particular future cost, must be considered very uncertain.

The only contract price for a large-scale electrolyser plant reported is by Nel for the delivery of 5,500 Nm³/hr (19.5 MW_{H₂HHV}) plant comprising 11 NEL A485 units to Tokuyama Corporation (see Figure 3.14). The order value was 160 MNOK (circa £15m) in 2013/14 equal to £760/kW_{H₂HHV}. This price most likely excludes installation, tie-in, electrolyser house, compressor, site preparation, etc.



Figure 3.14: Electrolyser house at Tokuyama Corporation facility.

The Felgenhauer and Hamacher study²⁰ from 2015 is based on 11 AEL and 5 PEM quotes for the purchase of commercial systems from 9 companies. This was intended for onsite hydrogen generation for logistic vehicles in South Carolina. Quoted AEL systems have electric power inputs of 0.35–3.35 MW whilst the electric power input for PEM is between 0.45 and 2.5 MW. The quotes include all necessary auxiliary components. The companies contacted were CETH2/Areva H2Gen, Hydrotechnik, Hydrogenics, ITM Power, McPhy Energy, Nel, Next Hydrogen, PERIC and Siemens. The performance data were anonymised to allow publication. The investment cost data from the suppliers were supplemented with estimates for shipping, foundation/shelter (if needed) and the ‘turnkey’ installation. The additional cost was estimated to account for 16% of the investment cost on the average for the most promising AEL technologies in terms of hydrogen production costs.

The hydrogen production rate is linearly dependent on the electrolysis cell area and it is mainly the cost of the auxiliaries that are reduced with increasing scale. This means the economy of scale effect is expected to be limited for multi-stack electrolysis systems compared to a single stack AEL electrolyzers basis, i.e. once MW scale is reached the economies of scale will diminish. This is shown in Figure 3.15. Hence, reaching larger stack sizes is significant.

²⁰ Felgenhauer, M. and Hamacher T, “State-of-the-art commercial electrolyzers and on-site hydrogen generation for logistic vehicles in South Carolina”; Int. J. Hyd. Energy 40, 2084–2090, 2015

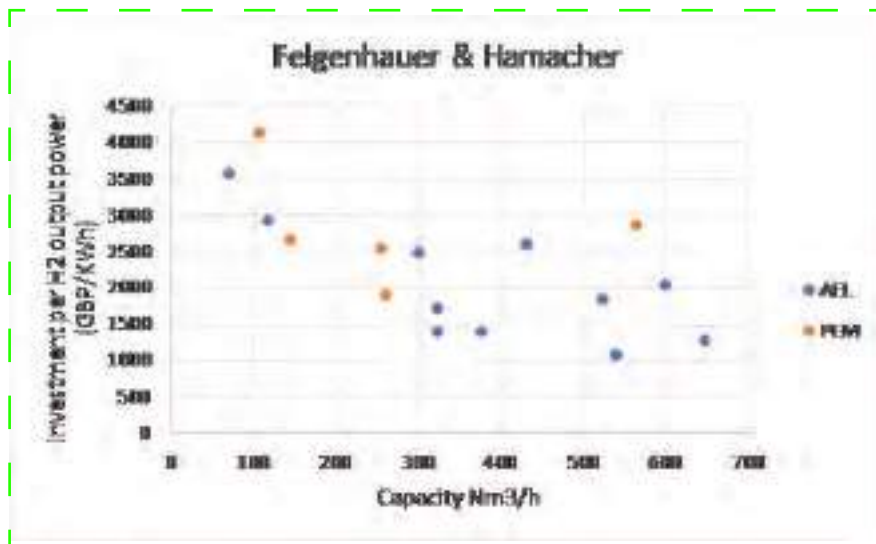


Figure 3.15: Capacity versus investment cost. Source: Felgenhauer and Hamacher

Table 3.4 summarises PEM and AEL investment cost data from literature. Data were converted to $\text{£/kW}_{\text{H}_2\text{HHV}}$ assuming a specific energy consumption of 5 kW/Nm^3 where necessary.

Study	AEL investment cost $\text{£/kW}_{\text{H}_2\text{HHV}}$		PEM investment cost $\text{£/kW}_{\text{H}_2\text{HHV}}$	
	Uninstalled	Installed	Uninstalled	Installed
Felgenhauer and Hamacher		1,000-1,300		2,000-3,000
Buttler and Spliethoff	900-1,600	1,000-1,800	1,600-1,900	
Götz et al	1,000-1,600		2,400-2,900	
Schmidt et al.	1,200		2,500	

Table 3.4: Water electrolysis investment cost from literature

AEL has the lowest investment cost with the largest and cheapest systems around $\text{£1,000-1,300/kW}_{\text{H}_2\text{HHV}}$. The largest PEM system is at $\text{£3,000/kW}_{\text{H}_2\text{HHV}}$, whereas the cheapest is at $\text{£2000/kW}_{\text{H}_2\text{HHV}}$. The authors consider that lower cost offers for PEM could be seen soon based on announcements from Hydrogenics, Siemens and Proton Onsite. The economic comparison is done at an output pressure of 10 bar (except for three atmospheric AEL electrolyzers) which are in the higher AEL cost range, starting at $\text{£2,900/kW}_{\text{H}_2\text{HHV}}$. This may be due to the less compact design.

Buttler and Spliethoff estimates the installed cost based on data from Felgenhauer and Hamacher and four earlier papers (including Smolinka (2011) and Bertuccioli (2014)). All references give cost per kW and uninstalled cost.

Here the data has been converted by assuming a specific energy consumption of 5 kW/Nm^3 of hydrogen. Smolinka estimates the installation cost to be 10% of the investment. For AEL, Buttler and Spliethoff reference installed cost in a range between £1,000 and $\text{£1,800/kW}_{\text{H}_2\text{HHV}}$, whereas the uninstalled cost is between £900 and $\text{£1,600/kW}_{\text{H}_2\text{HHV}}$. The PEM cost is almost twice as high with uninstalled cost ranging from £1,600 to $\text{£1,900/kW}_{\text{H}_2\text{HHV}}$.

The investment cost of Götz et al (2016) is based on other references (mainly German and French) and are presumably uninstalled cost²¹. The AEL cost is between £1,000 and $\text{£1,600/kW}_{\text{H}_2\text{HHV}}$ and the PEM cost is between £2,400 and $\text{£2,900/kW}_{\text{H}_2\text{HHV}}$. Götz refers a claim from Siemens that PEM CAPEX will be on par with AEL in 2018 but the source of this information was not found.

²¹ M. Götz et al., Renewable Power-to-gas: A technological and economic review, Renewable Energy 85 (2016) 1371-1390

3.3.5 1.5GW electrolyser based hydrogen facility

Based on the analysis above, a 1.5GW alkaline electrolyser-based hydrogen facility was designed and evaluated for the H21 NoE study. The facility was based on the Nel A-485 technology, which is the most referenced and proven electrolyser technology.

Key design parameters are given in [Table 3.5](#).

	UNITS	VALUE
Cell stack efficiency	kWh/Nm ³	3.8
Cell stack annual degradation	%	1
Stack lifetime	Years	10
Cell stack efficiency end-of-life	kWh/Nm ³	4.35
Cell stack efficiency average	kWh/Nm ³	4.08
Compression 0-80bar	kWh/Nm ³	0.23
Utilities and offsite	kWh/Nm ³	0.60
HVAC-DC losses	%	2
Average system efficiency	kWh/Nm ³	5.01
	kWh/kWh _{H2HHV}	1.41
	% (HHV)	70.9
End-of-life system efficiency	kWh/Nm ³	5.29
	kWh/kWh _{H2HHV}	1.49
	% (HHV)	67.1
Hydrogen production	MW _{H2HHV}	1,500
Average power consumption	MW	2,116
Water consumption	m ³ /h	380.3
Cooling water	m ³ /h	5,000

Table 3.5: Key design parameters for a hydrogen facility based on water electrolysis

The facility included the following units:

- 872 NEL A485 electrolyser units placed in 27 electrolyser houses with 32 units in each house – the transmitter function is included in the substation
- 2.2 GW HVAC to DC substation
- Desalination unit and cooling water system
- Hydrogen compression from 0 to 80 bar

Each electrolyser house would accommodate 8 x 4 A485 units. Each group of four electrolysers is equipped with common oxygen and hydrogen separators. Sufficient space is needed inside the electrolyser house for electrolyser replacement and maintenance.

The electrolyser house is equipped with detectors and necessary ventilation. The hydrogen scrubber and electrolyte tank are installed outside the electrolyser house. The entire 1.5GW facility comprises a total of 27 electrolyser houses. The facility requires a total area of 500,000 m² and the plot plan is shown in [Figure 3.16](#).

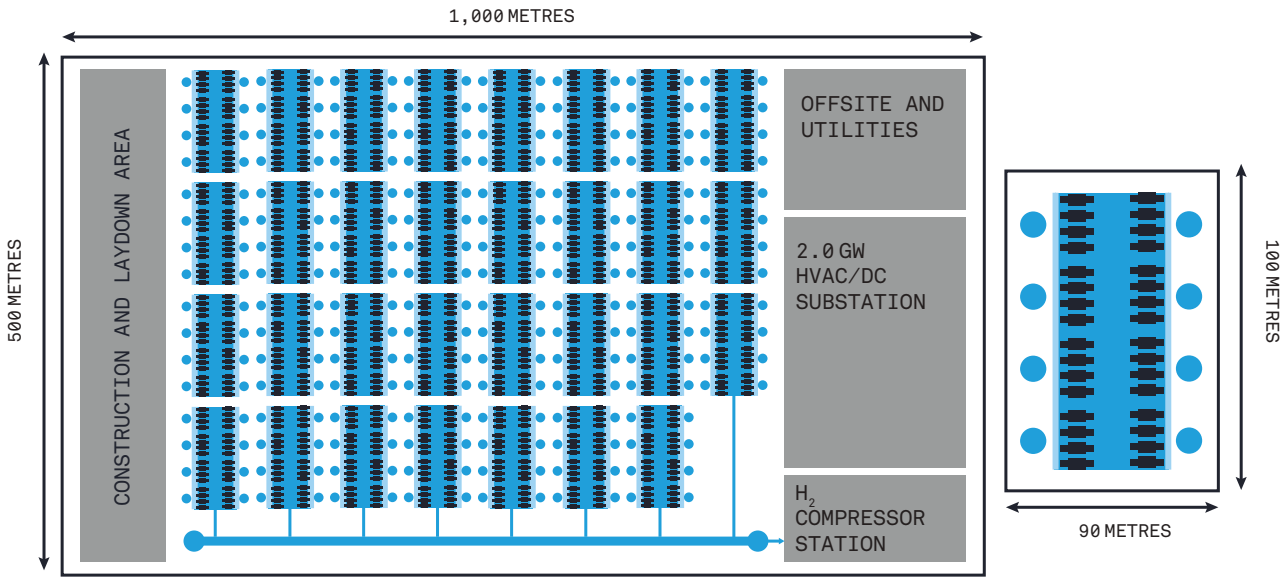


Figure 3.16: 1.5GW electrolyser facility plot plan

A cost estimate is established and shown in Table 3.6 below. All units outside the electrolyser plant are estimated using the same methodology as for natural gas reforming and ammonia.

The owners cost includes electrolyser house, connection to the grid, water intake to desalinisation unit etc.

Construction cost also includes site preparation for the electrolyser units and tie-in with remaining units, e.g. hydrogen to compressor and cables from the substation to the electrolyser units

The total cost is estimated at £1,740m equal to £1,160/kW_{H₂HHV}, which is the lower end of published cost data for AEL systems.

£M	ELECTROLYSER UNIT	OFFSITE, SUBSTATION, COMPRESSOR	TOTAL
Equipment	870	110	980
Bulk		95	95
Indirects		22	22
Construction		192	192
Home office		20	20
CMT		24	24
Other		6	6
EPC (£m)	870	469	1,339
Owners cost	80	50	130
Project management	20	23	43
Insurances	30	14	44
Contingency	90	94	184
TOTAL PROJECT COST	1,090	650	1,740

Table 3.6: CAPEX breakdown of a 1.5GW hydrogen plant based on alkaline water electrolysis

3.3.6 Operating cost

According to Ursua AEL electrolyzers are silent, reliable devices with little maintenance due to little or no rotational equipment.

Bertuccioli reports OPEX values of between 2 and 5% of CAPEX per year based on indicative figures from a few manufacturers willing to give such information. Felgenhauer and Hamacher reports annual O&M cost including maintenance/spare parts and operation/labour for the AEL of between £50 and £100/kWh_{H2HHV} and £100 and £180 /kWh_{H2HHV} for the PEM. The annual operating cost for this range is between 1.6 and 4.4% and 2.3 and 4.1% of the CAPEX for AEL and PEM respectively, it is uncertain whether these costs include the stack upgrades. For the 1.5 GW facility above this would result in an annual OPEX cost in the range of £27m to £77m per annum.

According to a 2018 publication by IRENA (International Renewable Energy Agency)²² the stack upgrade cost could be about 40-50% of system cost of an electrolyzer.

3.3.7 Safety

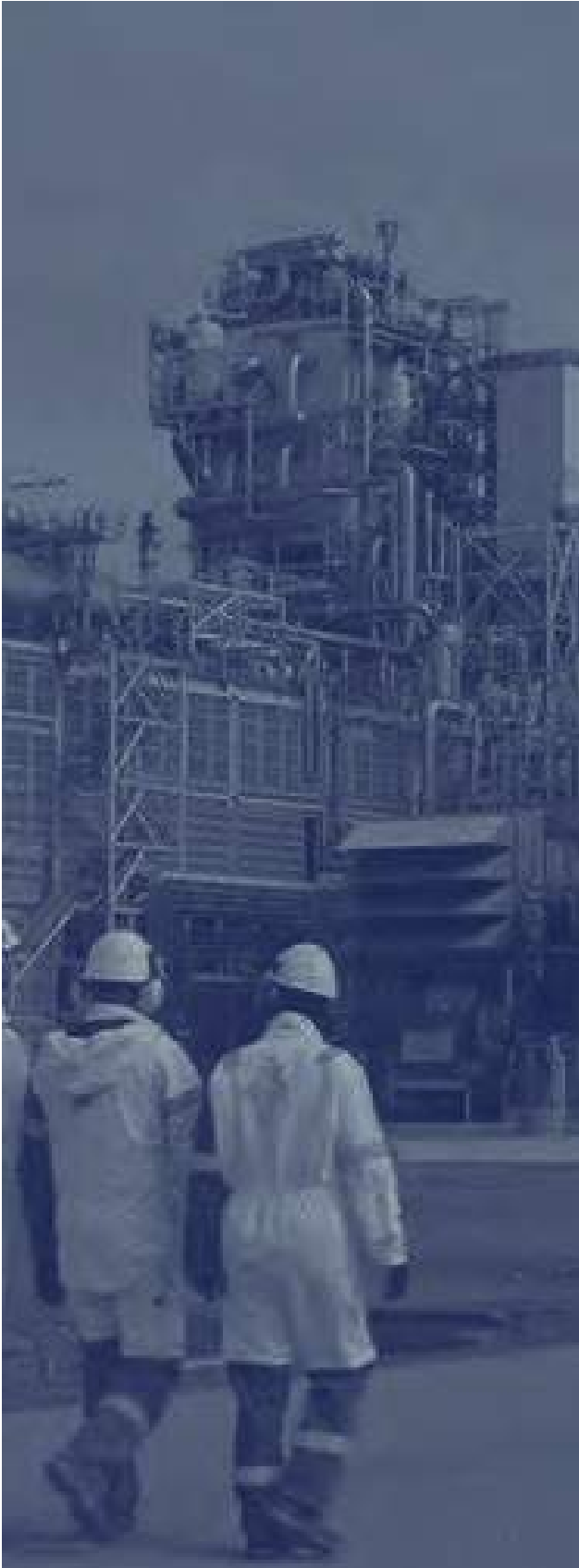
Large scale facilities will be installed in an electrolyzer house which increases risk in case of hydrogen leakage, requiring an early leakage detection system and good ventilation. The plants may need explosive limits.

No review was found that offers a systematic assessment of the safety issues. The only topic usually covered is the risk of formation of flammable gas mixtures inside the electrolyzers due to cross diffusion of hydrogen. The risk is highest for AEL due to a higher diffusion rate across the diaphragm. The minimum load limit is set to control this risk. A commercial AEL system will also have a safety shutdown system kicking in at 1-2% hydrogen in oxygen. Experience shows that this risk can be readily managed but is likely to be more challenging to control in a system with fast load changes which could mean the use of large-scale industrial batteries for electrolyzers of this scale become essential. It is worth noting that this risk is less for PEM but will increase with increasing operating pressure. Zeng and Zhang mention electrolyte leakage in AEL systems, causing corrosion in connections and seals, as a safety risk.

Carmo mentions two potential PEM safety risks. Firstly, titanium may self-ignite in pressurised oxygen. This can be avoided in a differential pressure configuration where oxygen is vented to the atmosphere. This will likely limit the maximum output pressure of hydrogen. Secondly, hydrogen embrittlement on the cathodic current collector and separator plates may cause hydrogen leakages.

Smolinka mentions increased mechanical stress caused by a stop/start operation regime impacting lifetime. This could potentially be a safety risk in pressurised systems.

Clearly, a more thorough and systematic assessment and comparison on basis of safety is required to mature the technology for large scale energy applications.



3.3.8 Water electrolysis summary

The analysis of AEL and PEM water electrolysis concludes that AEL is the preferred water electrolysis technology when assessed against applicability to a project like H21 NoE. The following key performance features for 1.5GW hydrogen production plant:

- Energy efficiency of 70.9% (HHV) average with 10 years between cell replacement or upgrade
- Capital cost of £1,160/kW_{H₂HHV} equal to £823/kW_{el} (£1,160/kW x 70.9% efficiency) for a fully functional hydrogen facility with all support systems
- Area need is 500,000m² (50 ha) equal to 33 ha/GW

Assuming a 24-month delivery time for an electrolyser unit it will take 4-5 years to deliver all the electrolysers for 1.5GW facility if production of 600 MW units annually is assumed (Thyssenkrupp). **Currently the supplier market does not have the capabilities to deliver 1-2 GW annually as required for a complete H21 NoE facility.**

Lifetime of cell stack and other equipment is critical since this can drive cost up. An intermittent power source can have a detrimental effect and is still an area with great uncertainty requiring future evaluation and testing.

3.4

Natural gas based hydrogen production

3.4.1 Methane reforming technology

A traditional hydrogen production plant consists of three main sections as illustrated in Figure 3.17. The different blocks will be described in more detail below.

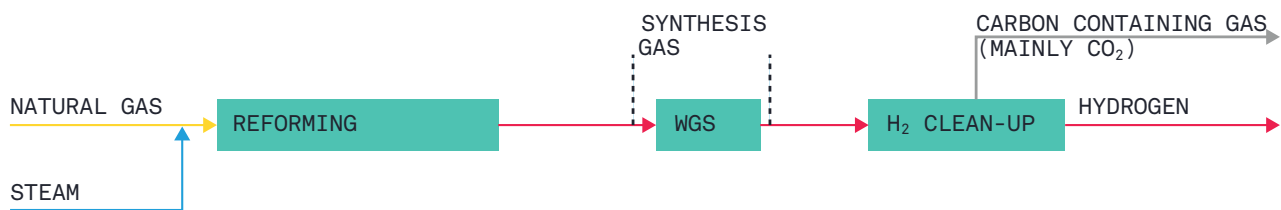


Figure 3.17: Simplified hydrogen production flow diagram

There are two different reforming technologies used to produce hydrogen gas from natural gas. They are Auto Thermal Reforming (ATR) and Steam Methane

Reforming (SMR). Both systems are appraised in this section for their overall applicability to the H21 NoE concept. The principals of the two reforming options are set out in Figure 3.18.

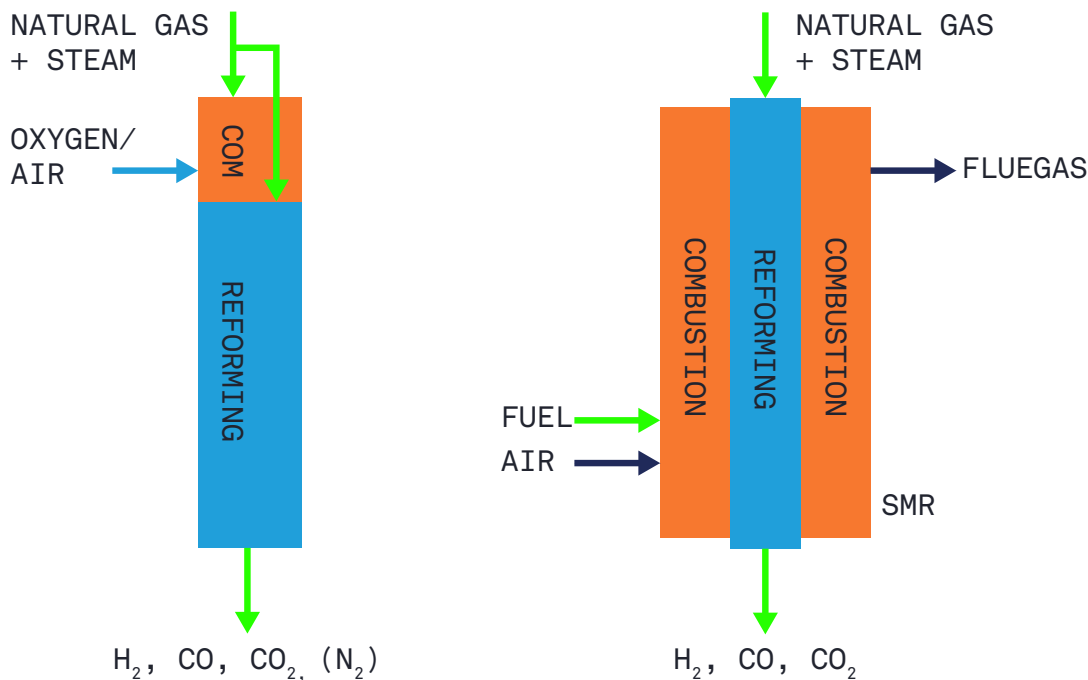


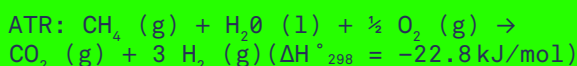
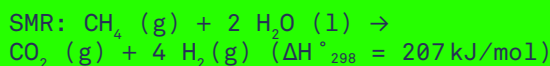
Figure 3.18: ATR and SMR simplified figures

The main difference between the two reforming technologies is how heat is supplied to the reforming reactions. The mixture of steam and natural gas must be heated to at least 900°C to be efficiently converted to H₂, CO and CO₂. The operating pressure can be in the range of 20 to 40 bar in an SMR and up to at least 60 bar in an ATR.

In the SMR reactor heat is transferred from a separate combustion section, whilst in the ATR reactor heat is generated by partial combustion of the reformer feed (natural gas) with oxygen. In practise, this means that a more CO₂ rich synthesis gas will be produced in the ATR than in the SMR since the combustion products (H₂O and CO₂) will end up in the synthesis gas. In an SMR the carbon containing fuel gas (mainly natural gas) is combusted with air in a separate combustion chamber and generated flue gas containing CO₂ will be emitted to the atmosphere in that case. This is an important difference between the two technologies regarding the potential to capture a major part of produced CO₂.

In both processes, ATR and SMR, the reformed gas at high temperature and pressure (i.e. the synthesis gas) will contain a significant amount of Carbon Monoxide (CO) which in an hydrogen process is being converted to CO₂ and more hydrogen at 200 to 400 °C in a Water Gas Shift (WGS) reactor section (see Figure 3.22). The cooling is mainly done by preheating water for high pressure steam generation and production of low pressure steam. The gas downstream of the WGS process contains mainly hydrogen (78%) and CO₂ (20%) with the remaining being unconverted methane and CO. CO₂ can then be removed in a H₂ clean-up section, liquified and pumped to a geological storage site (see Section 7.0). Clean hydrogen is then ready for use. If using a traditional amine CO₂ separation process more than 99.8% of the CO₂ can be removed from the shifted synthesis gas at this stage.

The overall reactions for the two reforming processes can be summarised the following way:



From a pure CH₄ to H₂ conversion perspective SMR is more efficient compared to ATR (you get more H₂ molecules per CH₄ molecule). However, as seen from the above overall reactions the SMR route requires energy (207 kJ / mol CH₄ equal 207,000 kJ/kmol / 4 (mol H₂/mol CH₄) / 3,600 kWh/kJ / 2.02 kg/kmol (MWH₂) = 7.12 kWh/kgH₂) which is equivalent to between 20 and 25% of the energy content of hydrogen. Whereas the ATR scheme is generating energy (22.8 kJ/mol CH₄ equal 22,800 kJ/kmol / 3 (mol H₂/mol CH₄) / 3,600 kWh/kJ / 2.02 kg/kmol (MWH₂) = 1.05 kWh/kgH₂) equal to between 2 and 3% of the energy content of hydrogen (you get approximately the equivalent energy in the form of H₂ in an ATR as with an SMR).

Separation of CO₂ from synthesis gas is a well proven technology which is used in most of the ammonia plants in the world. Due to the nature of the catalyst used in the ammonia synthesis reaction, only very low levels of oxygen-containing compounds (especially CO, CO₂ and H₂O) can be tolerated in the synthesis (hydrogen and nitrogen mixture) gas. Therefore, high conversion of methane to hydrogen is required as well as high CO₂ capture rates. Removing CO₂ down to less than 500 ppm in the hydrogen stream is a key design requirement in ammonia production.

As ammonia production consumes more than 50% of global hydrogen production (30Mt and 140GW hydrogen capacity) the hydrogen production part of the ammonia plant is therefore a perfect starting point for a hydrogen production plant which includes capture of CO₂ based on proven and references design at scale.

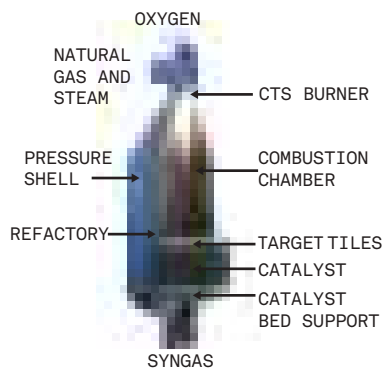
Each of the three blocks described in Figure 3.17 includes several process elements which are described in more detail below. In addition, different utilities will be needed, e.g. power generation, boiler feed water preparation, cooling systems, steam systems, waste water treatment, chemical storage, flare, air separation (only ATR), etc. is described in the following section.

Description of ATR and SMR

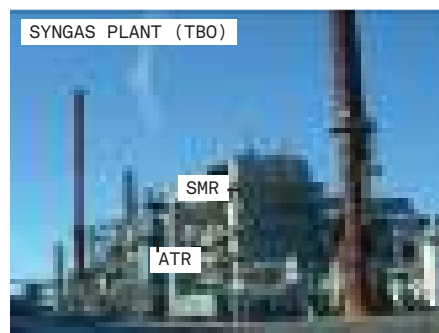
To allow production of hydrogen with minimum inerts in an ATR process, pure oxygen must be fed to the reactor. Oxygen blown ATR is a proven technology and is, for example, used in the production of methanol at Tjeldbergodden, Norway. Traditionally a combination of SMR and ATR is used in the production of methanol.

The largest oxygen blown ATR in operation is in the Oryx gas-to-liquids plant in Qatar and was delivered by Haldor Topsoe. The unit has been in operation since 2006 and produces 600,000Nm³/h of syngas equal to 2.1GW of hydrogen.

The SMR and ATR sections at Tjeldbergodden (TBO) methanol plant are illustrated in Figure 3.19. The H₂ production capacity is about 6.2 million sm³ hydrogen per day (about 0.9 GWh). One third is generated in the SMR and two thirds are generated in the ATR. The ATR reactor is significantly more compact than the SMR²³.



ATR REACTOR



PICTURE OF THE
COMBINED SMR AND ATR
SECTION TJELDBERGODDEN



SMR TUBULAR REACTOR

Figure 3.19: SMR and ATR installed at Tjeldbergodden methanol plant

The SMR reactor consists of a high number of 13-15 m long tubes filled with a catalyst. High conversion is thermodynamically favoured by low pressures, high steam to carbon (S/C) ratio and high temperatures. A high S/C ratio will also reduce the risk of carbon deposition on the catalyst surface but using more steam will reduce process efficiency. The SMR is disadvantaged by having lower methane conversion compared to an ATR with unconverted methane ranging from 5-10% depending on design and operating conditions. For hydrogen production where high CO₂ capture rates are required (above 90%) this can be a challenge.

An ATR reactor includes a fixed catalyst bed and can be operated at a higher temperature and pressure than a tubular reformer. This allows high conversion of methane and high synthesis gas pressure at a lower steam to carbon ratio. Reduction in the quantities of steam compared with an SMR may improve process efficiency. High synthesis gas pressure is an advantage because it produces hydrogen at high pressure reducing the need for hydrogen compression²⁴.

²³ A. Gedde-Dahl and K. Jørgen Kristiansen (Equinor) and H. Holm-Larsen (Haldor Topsøe), "The 2,400 MTPD Methanol Plant at Tjeldbergodden"

²⁴ E. Sorensen (Haldor Topsoe Inc.), "Autothermal Reforming for efficient and versatile syngas production"

Both SMR and ATR reactors are operated at a very high temperature and the gas streams downstream of the units can be more than 1,000°C. Thus, heat recovery can be very challenging. Normally, the heat is being recovered by producing steam at high pressure (e.g. 120 bar). This will keep the metal surface at reasonably low temperatures.

Some of this heat can be recovered using a heat exchange reformer or a Gas Heated Reformer (GHR), (the name used depends on the vendor) either in series or in parallel with the main reformer as illustrated in [Figure 3.20](#).

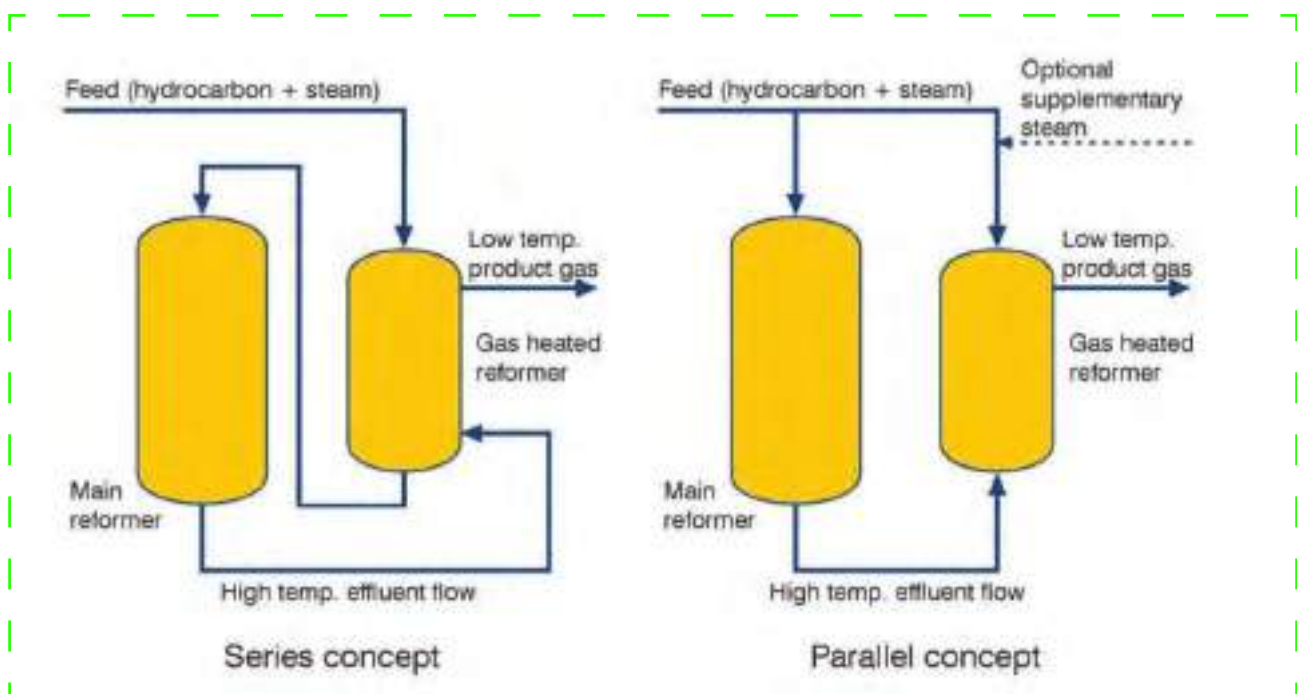


Figure 3.20: Illustration of a gas heated reformer used either in series or in parallel with the main reformer

The main advantage of installing a heat exchanger reformer is more efficient use of recovered heat. Recovered heat is used to convert more natural gas to hydrogen. Therefore, less fuel is needed to be combusted which reduces the natural gas consumption. In an ATR based process it will also reduce the consumption of pure oxygen by at least 15%.



In the series concept all the feed gas passes through the heat exchange reformer where it is heated and partly converted. The gas is then fed to the main reformer where final conversion takes place. The hot effluent from the reformer is cooled in the heat exchanger reformer and the remaining heat is used for reforming. The largest heat exchanger reformer in operation is about one tenth of the size required for between 1 and 2GW. Significant scale-up of this technology will therefore be needed.

In the parallel concept, the feed gas is split into two streams. One stream is sent to the main reformer where it is heated and converted. The other stream is sent to the heat exchange reformer. The main part of the heat for the heat exchange reformer is provided by cooling of the effluent from the main reformer. A part of the reaction heat will be provided by heat exchange of the reformed gas with the catalyst bed. The parallel heat exchange reformer technology is well demonstrated in parallel with an air-blown ATR in an ammonia flow scheme, but less proven in combination with an oxygen-blown ATR.

A heat exchanger reformer is, in most cases, used in combination with an ATR reactor but can also be used in combination with an SMR. An example of a gas heated reformer (GHR) applied in parallel with the ATR is KBR's so-called Kellogg's Reforming Exchanger System (KRES) reformer. The tubes in the KBR reforming exchanger are open-ended and hang from a single tube sheet at the inlet cold end to minimise expansion problems. The recommended ATR technology in this report is based on parallel heat exchanger reforming which is referenced at scale. Other suppliers of gas heated reformers are Johnson-Matthey and Haldor Topsoe.

The main challenge with both a steam boiler and particularly a gas heated reformer is the control of metal dusting corrosion. Metal dusting is an attack on the metal surface resulting in pits or loss of material. The mechanism is related to the destruction of the protective oxide layer on the metal surface. This is followed by carbide formation in grain boundaries causing disintegration of the metal. The potential for metal dusting is highest in Carbon Monoxide (CO) rich gases at temperatures of 400-800°C. These are conditions which are likely to prevail for equipment during cooling of the synthesis gas downstream of the ATR or SMR.



Description of natural gas pre-treatment and reforming

Treatment of the natural gas feed up-stream of the ATR or SMR unit will be needed. In most cases the natural gas feed contains minor amounts of sulphur that must be removed because it will poison the different reforming and WGS catalysts in the downstream process. Any organic sulphur in the natural gas feed is hydrogenated in a catalytic reactor (to H_2S) by mixing in a small amount of recycled hydrogen (Figure 3.21).

H_2S is then adsorbed in a zinc oxide bed and reacts to form zinc sulphide, which is removed as a solid waste. The desulphurisation operating temperature is between 260-430°C and the pressure can be up to at least 60 bar. Sulphur poisoning may be permanent and pre-treatment through desulphurisation is therefore crucial for successful production of hydrogen from natural gas.

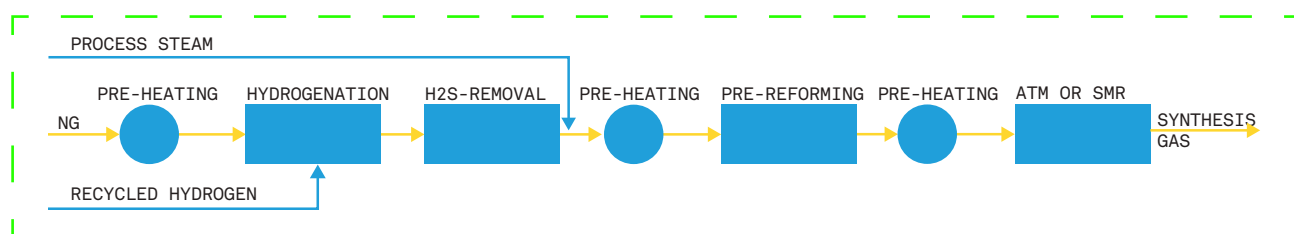


Figure 3.21: Reforming front section

In hydrogen plants using natural gas containing heavier hydrocarbons like ethane, propane and butane, an **adiabatic pre-reformer** can be included upstream of the SMR or ATR. Adding steam to the gas will convert these compounds to hydrogen, CO and CO₂ at 450-500°C. Since this is an endothermic reaction the temperature will drop. This will both allow adding more energy to the gas and allow pre-heating the gas inlet to the reformer unit to an even higher temperature eliminating coke formation. In an SMR, this will reduce the required reformer tube area and therefore reduced cost. In an ATR based process the advantage is primarily reduced oxygen consumption.

Heating of the gas inlet in the different reactors is done using hot process gas, steam, flue gas streams or by use of a fired heater. This depends on the selected reformer technology. If SMR is used, then most of the preheating can be done using hot flue gas exiting from the combustion section

An ATR reactor produces no flue gas so an additional fired heater must be installed to heat the feed gases and for plant start up. The heater and boilers in an ATR based hydrogen plant will be fuelled by a mixture of 90% hydrogen and 10% natural gas, the 10% natural gas is added to maintain flame stability. This results in some CO₂ emissions affecting overall CO₂ capture rate. In a future Front-End Engineering Design study (FEED, see Section 10.0) 100% hydrogen combustion will be examined to achieve even higher CO₂ capture rates.

Description of Water Gas Shift (WGS)

Downstream of the reforming process, the syngas undergoes water gas shift where CO is reacted with water (i.e. steam) over a catalyst to produce additional hydrogen and CO₂ (Figure 3.22).

The overall reactions for the water gas shift can be summarised as follows:

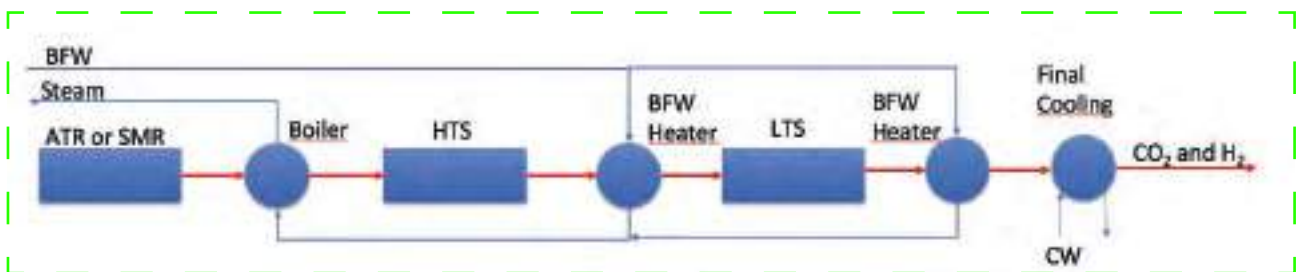


Figure 3.22: WGS Section including steam generation

The process can be done in two steps if high conversion of CO is needed; high temperature shift (HTS) and a low temperature shift (LTS). In the HTS, the mole fraction of CO is reduced from typically between 10 to 13% to between 2 to 3%, with an inlet temperature between 350-400°C. In the LTS, the CO concentration is further reduced to between 0.2 and 0.4% at temperatures of between 190 and 250°C. The lower temperature limit is set due to the water dew point of the gas which is a function of pressure and steam content. Any condensation could damage the catalysts in the WGS chamber. Finally, the hydrogen and CO₂ containing gas is cooled to about 40°C resulting in all water being condensed and separated.



Description of H₂ clean-up

The synthesis gas leaving the WGS Section contains mainly hydrogen and CO₂; however unconverted CO and methane as well as argon and nitrogen will also be present. High amounts of methane will be present in the synthesis gas for an SMR based hydrogen production scheme.

The design and selection of the hydrogen clean-up method is based on the required hydrogen specification. In the case of fuel cells, hydrogen purity above 99.999% is presently required. Whereas for combustion in boilers and gas turbine operation, 98% purity or lower is required.

In traditional large-scale hydrogen production with SMR with a requirement for hydrogen above 99% purity, the preferred choice is Pressure Swing Adsorption (PSA)

The PSA system adsorbs the CO, CO₂, and other impurities in a fixed-bed adsorber. The adsorbent does not adsorb the hydrogen. Apart from the pure hydrogen product, the PSA system produces a low-pressure gas stream (the tail gas). It contains all the impurities present in the feed gas and hydrogen used for regeneration of the adsorbent. 10-15% of the hydrogen leaves with the tail gas.

Normally the PSA tail gas is used as fuel gas for the steam reformer furnace where it provides at least 50% of the total heat input depending on the process conditions. Therefore, all generated CO₂ will end up in the flue gas and will be emitted to the atmosphere. Using this technology makes the CO₂ capture more challenging since CO₂ must then be removed from the flue gas at low pressure.



Figure 3.23: PSA system

Pressure Swing Adsorption (PSA) can also be used to separate out CO₂ and unconverted gas. Molecules have different affinities for various types of adsorbents and this phenomenon can be utilised to purify gases.

Since the hydrogen for H21 NoE will be used in heaters and boilers the need for a PSA is not required. Sufficient clean up can be achieved by the CO₂ capture technology and by selecting optimised design parameters.



Description of the Air Separation Unit (ASU)

To allow production of pure hydrogen in an ATR process, pure oxygen must be fed to the reactor.



Figure 3.24: ASU plant at Tjeldbergodden (TBO) in Norway

Traditional cryogenic air separation is a mature technology with few possible future improvements. The short-term direction of ASUs is increased single unit capacity and process optimisations depending on possible integration with O_2 consuming processes. The ASU at Tjeldbergodden methanol plant in Norway (1,200 t O_2 /day) is shown in Figure 3.24. A 2,000 m³ buffer storage for Liquid Oxygen (LOX) is installed to help maintain a high on-stream factor.

Filtered and purified air is compressed to between 5 and 6 bar and cooled to minus 180°C. Then it is expanded and cooled to below the boiling point of oxygen (-183°C). As a result, it is partially liquified. By means of a separation column, the oxygen and nitrogen are separated. Gaseous nitrogen collects at the top of the separation column and liquid oxygen collects at the bottom of the column. Used in a ATR based hydrogen plant, liquid oxygen is pumped to required pressure and evaporated.

3.4.2 CO₂ capture technology

Converting gas networks to 100% hydrogen will facilitate bulk availability of a fuel, hydrogen, which has close to zero emissions at point of use. However, as with electrolysis, the net emissions of the systems producing the hydrogen need to be considered to ensure such a strategy meets UK climate change obligations. For hydrogen produced via natural gas using either SMR or ATR technology, the carbon must be captured and stored indefinitely to achieve the overall low carbon objectives. This requires two areas of consideration. Firstly, the total percentage carbon that can be captured during the production process and secondly, the availability of storage for permanent sequestration.

The CO₂ is produced from the following processes:

- During the steam methane reforming and water-gas shift reactions
- During the combustion of carbon containing fuel gas in the SMR, ATR pre-heater or from any auxiliary boilers and power generation unit using a carbon containing fuel

The realistic percentages of carbon that can be captured during the SMR/ATR process are set out below.

Capturing a major part of generated CO₂ requires additional process steps. The main additional process steps include CO₂ absorption process and a CO₂ compression unit or a CO₂ liquefaction, depending on if CO₂ is going to be transported in a pipeline to the storage site or transported by ship.

There are three CO₂ capture alternatives:

- Capture of CO₂ from produced synthesis gas at high pressure using, for example, activated Methyl DiEthanol Amine (aMDEA) solvent
- Capture of CO₂ from generated flue gas at low pressure using Mono Ethanol Amine (MEA) solvent or similar
- Capture of CO₂ from both the synthesis gas and from flue gas

High pressure CO₂ capture

To achieve deep CO₂ removal of more than 99%, a mixture of water (circa 50wt%), MDEA and piperazine (4-6wt%) is frequently used. This solution (or solvent) is called activated MDEA, (i.e. aMDEA).

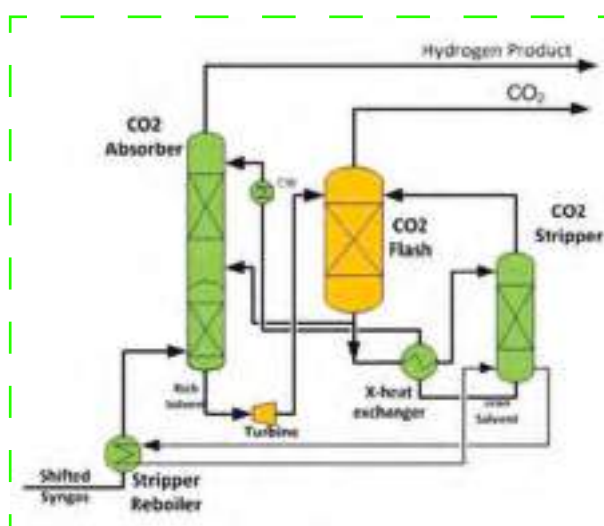


Figure 3.25: High pressure aMDEA CO₂ removal

The principle of CO₂ absorption at high pressure is that pressurised CO₂ containing gas passes through a reactor that contains the alkaline aMDEA scrubbing solution. This results in an acid-base reaction. CO₂ reacts with aMDEA and stays in the absorbent solution. The absorbent solution pressure is then reduced to say between 1 and 2 bar and some CO₂ will be flashed off (evaporated). One part of the solvent (semi-lean) is recycled to the absorber. The other part is heated in another reactor, (i.e. stripper) to release more CO₂. This purified aMDEA solution (containing very little CO₂) is then recycled to the top of the absorber and used again. The semi-lean absorbent removes the bulk of CO₂ while the purified solvent is removing the remaining CO₂ left in the gas (Figure 3.25).



If hydrogen is being produced using an SMR reactor, at least 35% of the natural gas feed is combusted at low pressure to generate heat for the steam methane reforming reactions. The resulting CO₂ containing flue gas is normally emitted to the atmosphere. Around 60% of the carbon fed to the hydrogen process can be separated as CO₂ using the aMDEA process in an SMR based H₂ process. If the produced hydrogen was used as fuel for SMR all CO₂ would be captured at high pressure. This option was examined but rejected since almost 30-40% of the produced hydrogen would be used to fuel the SMR.

If hydrogen is being produced using an oxygen fired ATR process, most of the carbon will end up in the syngas. Some fuel is, however, needed for preheating purposes and for generating steam during start-up and shut down. This fuel can be either natural gas or produced hydrogen. Either way more than 90% of the carbon in the natural gas fuel can be separated out as CO₂ using the aMDEA process in an ATR based H₂ process.

In this study, the target is to capture close to 95% of all carbon used in the production of hydrogen and the conclusion is that the ATR process is an attractive concept compared to the SMR route. It is, in principle, possible to achieve high CO₂ capture rates via the SMR route, but this would be traded-off with lower energy efficiency and higher cost.

More than 60% CO₂ can also be captured in an SMR based process by adding a reformer flue gas CO₂ capture unit (see Figure 3.32). Due to the low pressure of CO₂ in the reformer flue gas, a different CO₂ capture process must be used.

Low pressure (flue gas) CO₂ capture

The SMR flue gas CO₂ capture option requires a technology element not historically used in traditional hydrogen production plants due to no previous requirement to capture the CO₂. Flue gas CO₂ capture is a well-established technology and was developed mainly to capture CO₂ from coal or natural gas fired power plants but also from industrial flue gases. The common challenge in this case is the low partial pressure of CO₂ that require a very active absorbent (a medium to combine with the CO₂ in the flue gas and prevent it escaping to atmosphere), for example Mono Ethanol Amine (MEA). It is also called a solvent or solution since it contains a mixture of MEA and water (60-70%) and optionally other amines. The drawback is that this solvent needs a significant amount of energy to remove the CO₂ from it. This is done in a stripper column whereby steam promotes desorption (release) of the CO₂ from the solvent, at about 120°C, to be subsequently transported and captured in the appropriate carbon capture utilisation on/or storage system. Up to about 90% of CO₂ in a flue gas can be captured but this process will need about four times more steam (energy) per kg captured CO₂ than if CO₂ is captured at high pressure with aMDEA.

This low-pressure solvent-based CO₂ capture technology is illustrated in Figure 3.26. It has been demonstrated at Technology Centre Mongstad (TCM) which is a joint venture between the Norwegian state, Equinor, Shell and Total (see Figure 3.27).

For the H21 NoE project the options with a potential high CO₂ capture rate greater than 90% are set out in detail in the following section.



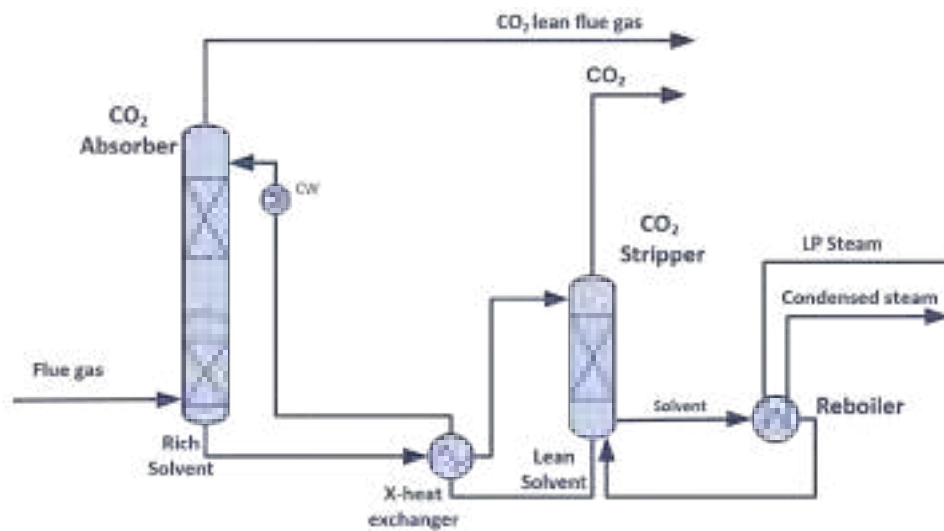


Figure 3.26: Low pressure CO₂ capture



Figure 3.27: Amine test plant at TCM

3.4.3 Optimised CO₂ capture solutions for ATR and SMR hydrogen production

To determine a realistic and credible carbon capture percentage from the ATR/SMR technology the study needed to focus on a specific capacity production plant.

1.5GW was selected as the appropriate capacity for modular scale up in the context of H21 NoE. This section is highly detailed, but it is important for full transparency as the carbon capture rates for the system are critical to the overall strategy for UK Government.

Design parameters for system analysis

- Plant capacity (maximum train capacity):
 - Export of 1.5GW hydrogen
 - Carbon capture rate: Target is 95% (this is based on the overall system including natural gas feed for H₂ production, natural gas fuel for the process, (i.e. efficiency loss), direct emissions from the hydrogen facility, unconverted CO and methane in the hydrogen product, fuel used in electric power production for the plant and compressors etc.)
- It is assumed that required electric power can be imported from the grid
- Natural gas composition used in this study (Table 3.7).

COMPONENTS	DRY MOL %	COMPONENTS	DRY MOL %
N ₂	0.77	i-C ₄ H ₁₀	0.05
CO ₂	2.10	n-C ₄ H ₁₀	0.08
CH ₄	89.47	C ₅ +	0.04
C ₂ H ₆	6.61	H ₂ S	5 mg/ Nm ³
C ₃ H ₈	0.87	HHV(MJ/Nm ³)	41.255

Table 3.7: Natural gas composition

- Ambient conditions
 - Barometric pressure: 760 mmHg (a)
 - Design air conditions: 22°C and 85% relative humidity (used for air coolers, fans and compressors)
- Fresh water:
 - Supply temperature: 16°C
 - Supply pressure: 8 bar
- Hydrogen product
- H₂ purity: > 97.5%
- Export pressure: 80 bar
- CO₂ product
 - CO₂ purity: > 98%
 - Water content: max 100 ppm
 - Export pressure: 300 bar

Calculation definitions

The definitions for the carbon capture and energy efficiency calculations are as follows:

Carbon Capture = (Carbon in CO₂ Product)/(carbon in NG Feed + Carbon in NG Fuel)

Energy Efficiency=(Energy in H₂ Export Stream (MW))/(Energy consumed for H₂ production (MW) + Electrical Power Import (MW))

Energy efficiency is based on HHV for NG and H₂ streams. Note HHV is the higher heating value which is the total energy contained in the fuel, LHV or Lower Heating value is the HHV minus the latent heat of vaporisation, i.e. energy used vaporising water during combustion.

HHV is an appropriate metric to use in the SMR/ATR energy consumption as part of the water condensation heat is reused in the process. LHV is appropriate to use in electric generation as water condensation heat cannot be reused in the production process.

Carbon footprint

The emission rates are for a plant exporting 1.5GW H₂.

The carbon footprint is estimated as follows (“ATR + GHR HP CC” as an example):

- The carbon footprint of natural gas feedstock: 185.3 kg/MWh
- The carbon footprint of produced hydrogen at an estimated carbon capture rate (94.12%) and efficiency (exclusive of power import) of say 83.1%: would be (185.3 ÷ 0.831) * (1-0.9412) = 13.1 kg/MWh (N.B. When reading Section 4.0 the carbon footprint is 14.4 kg/MWh as the power in the H21 NoE design concept is generated internally marginally increasing the carbon footprint of the system)



Modelling

Heat and mass balances for the different options are estimated using UniSim Design Suite from Honeywell.

UniSim Design Suite is a recognised process modelling software (preferred process simulation tool in Equinor) that provides steady state and dynamic process simulation in an integrated environment. The output from this tool is heat and mass balances and equipment design data that can be used to produce specified equipment lists needed to estimate the cost of the total process.

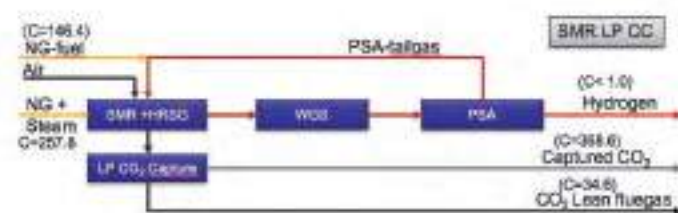
Well documented thermodynamic property databases and unit operation models are used to estimate accurate process data. UniSim Design includes over fifty-unit operations representing process equipment (i.e. separators, heat exchangers, distillation columns, reactors etc.).

Process modelling tools like UniSim are widely used in the process/chemical and oil and gas industry in design, engineering and operations.

CO₂ capture from an SMR based hydrogen process – 1.5GW H₂ production

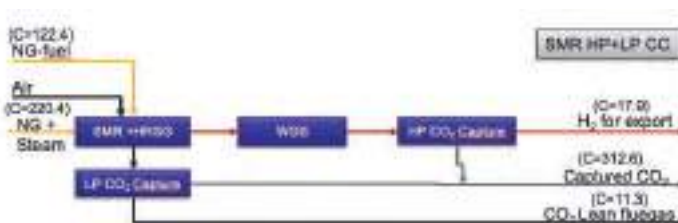
There are multiple alternatives for SMRs which can achieve high capture rates but with very different efficiency. So, the reader can easily see the difference, two options representing cases with either low or high efficiency are presented diagrammatically in Figure 3.28 with a brief description of the two alternatives in the following figures²⁵.

SMR OPTION 1: SMR AND PSA PURIFICATION AND LOW PRESSURE CO₂ CAPTURE (SMR LP CC)



Option 1: 1.5 GW H ₂ export	
C-capture (%)	91.2
CO ₂ -footprint (g/kWh)	23.8
Efficiency (%-HHV)	68.8
Power Import (MW)	0

SMR OPTION 2: SMR WITH BOTH SYNGAS HIGH PRESSURE CO₂ CAPTURE AND FLUE GAS LOW PRESSURE CO₂ CAPTURE (SMR HP+LP CC)



Option 2: 1.5 GW H ₂ export	
C-capture (%)	91.2
CO ₂ -footprint (g/kWh)	20.1
Efficiency (%-HHV)	79.5
Power import (MW)	35.6

Figure 3.28: Schematics of the two SMR concepts including the carbon mass balance (CO₂ losses in condensates around 1t/h)

²⁵ C-Capture (%) / Efficiency from UniSim Output, CO₂ Footprint based on NG parameters in Table 3.7. C= CO₂ equivalents in stream (t CO₂/h). Power consumption (100%) included in efficiency numbers.

SMR Option 1: SMR and PSA purification and low-pressure CO₂ capture (SMR LP CC)

This option is based on the same principles as the SMR described in the original H21 Leeds City Gate (LCG) report. In this type of SMR plant, all the CO₂ is emitted from the flue gas of the steam reformer at close to atmospheric pressure. This is because predominantly all CO₂ is removed by the PSA unit and is recycled to the reformer combustion section. This includes unconverted methane and CO and at least 10% of the hydrogen product used to purge the PSA unit.

Heat is recovered in a Heat Recovery Steam Generation (HRSG) unit. The HRSG unit also contains any auxiliary boiler systems needed to operate the plant.

Since CO₂ must be captured from the SMR flue gas at low pressure a significant amount of low-pressure steam (around 4 bar) must be used to capture the CO₂ (see SMR flue gas – low pressure CO₂ capture process description in the previous section). In this case much more steam than can be generated within the H₂ plant will be needed. Therefore, a supplementary boiler must be included to generate enough steam thereby increasing the fuel consumption significantly.

It is advantageous to generate steam at high pressure and reduce the steam pressure to 4.4 bar in a power generating steam turbine to generate low pressure steam for the capture plant. The steam turbine in this case will generate all the power needed for the hydrogen process including the CO₂ capture process and compression of CO₂ and hydrogen. Therefore, no power import is needed in this case. Since the PSA system recycles about 10% as reformer fuel (reducing the H₂ export), the SMR capacity also must be increased by about 10%. This, again, will increase fuel consumption and the amount of CO₂ that must be captured. Due to the high natural gas consumption (mainly due to the production of steam for the low-pressure CO₂ capture plant), the overall efficiency will be low – approximately 69% (HHV).

SMR Option 2: SMR with both syngas high pressure CO₂ capture and flue gas low pressure CO₂ capture (SMR HP+LP CC)

A second SMR alternative is to capture CO₂ both from the high-pressure syngas and from the reformer flue gas. The disadvantage is that we will need two different CO₂ capture units increasing CAPEX. However, energy needed to capture CO₂ at high pressure is only 25% of the energy needed to capture CO₂ at low pressure.

This concept is more energy efficient than SMR option 1. The main reason is that waste heat from final cooling of the reformer syngas can be used to remove most of the CO₂ in the high-pressure aMDEA system. Since the amount of CO₂ to be removed from the reformer flue gas is only 35% of the total amount (the rest has been removed in the high-pressure CO₂ capture system as part of the syngas processing), sufficient amounts of low pressure steam are available from the H₂ plant for removing the CO₂ from the flue gas. Therefore, less fuel will be needed compared with option 1, but electric power must be imported in this case.

High pressure steam at 128 bar is generated in the H₂ plant for both options. The steam is first used to generate power in an efficient high-pressure steam turbine. The resulting medium pressure steam at about 59 bar (pressure is reduced through the turbine) is split in two streams. About 67% is used as process steam and mixed with the natural gas in the reformer. Excess steam is used to generate power in a medium pressure steam turbine which will generate low pressure steam at 4.4 bar. This low-pressure steam is then used to remove CO₂ from the solvent in the low-pressure CO₂ absorption plant.

With an overall carbon capture rate of 91%, a plant efficiency of 79% will be possible. This is significantly higher than option 1.

If a Pressure Swing Adsorption (PSA) unit is included producing 99.99%+ pure hydrogen, up to 96.9% CO₂ can be captured as the PSA tail gas (containing a high content of hydrogen) is used as fuel in the reformer. This will, however, reduce the efficiency to about 75% (HHV) and is over engineered for gas network delivered hydrogen as will be required for the H21 NoE conversion. The same applies to option 1 where using a PSA will produce a very clean hydrogen gas.

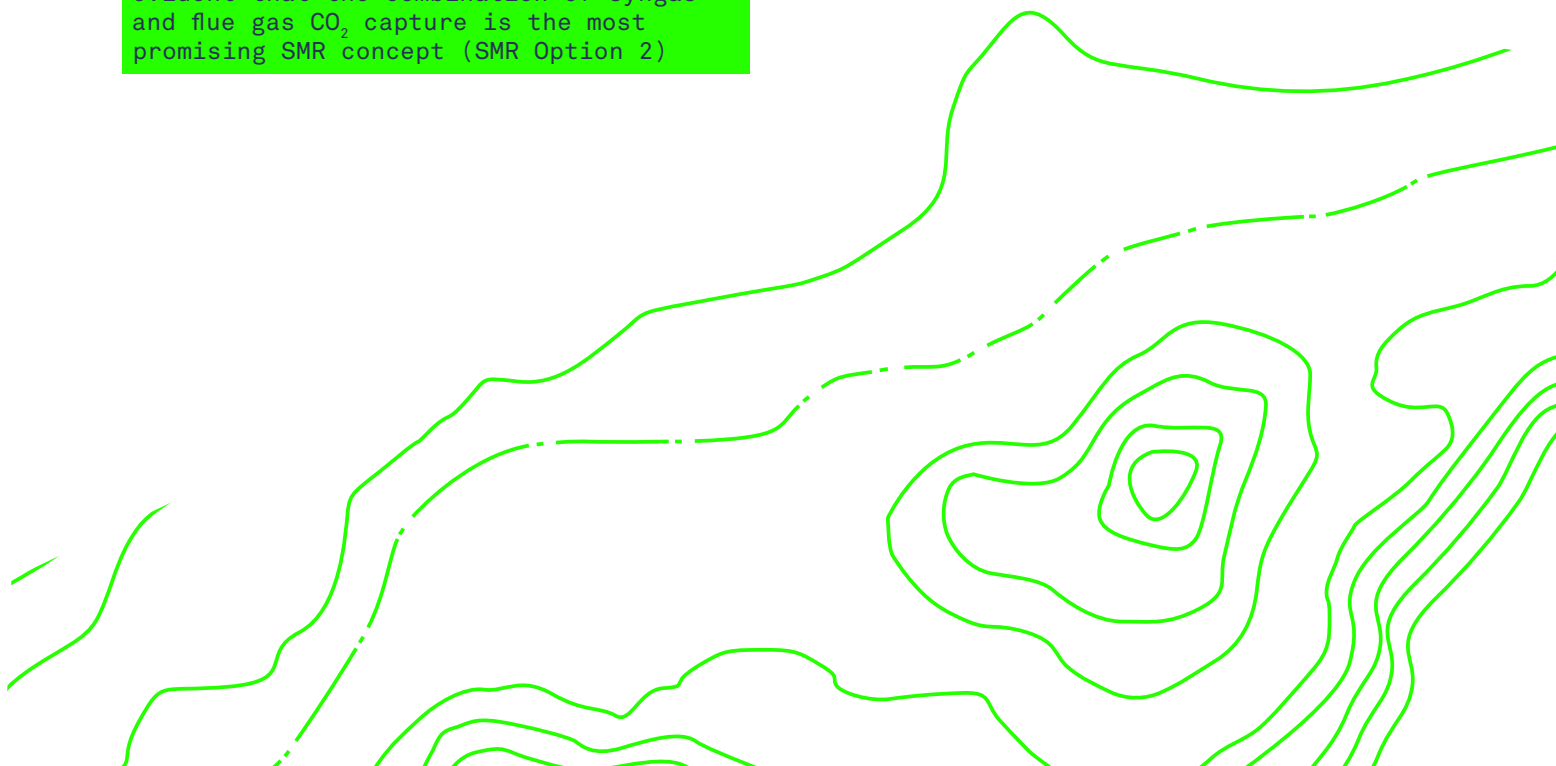
Since a low-pressure CO₂ capture plant is included in option 2, a combined cycle gas turbine (CCGT) can be installed to generate electric power in lieu of importing electric power from a hydrogen fired power plant. This would require capturing CO₂ from the gas turbine flue gas in the CO₂ capture facility. Both CO₂ capture rate and efficiency will be about the same as for option 2. This makes this process more flexible regarding power supply.

The overall performance of the two SMR options are summarised in [Table 3.8](#).

1.5GW H ₂ EXPORT	SMR OPTION 1 SMR LP CC	SMR OPTION 2 SMR HP+LP CC
CO ₂ capture (%)	91.1	91.2
CO ₂ footprint (g/kWh)	23.8	20.5
Efficiency (%-HHV)	68.8	79.5
NG consumption (MW)	2,182	1,850
Electric power import (MW)	0	35.6
Steam production (t/h)	844	496
CO ₂ export (t/h)	369	312
H ₂ purity (%)	97.5	97.5
H ₂ export pressure (bar)	80	80
Number of SMR's	2	2

Table 3.8: Comparison of SMR alternatives

Since the target is high process efficiency and low carbon footprint it is evident that the combination of syngas and flue gas CO₂ capture is the most promising SMR concept (SMR Option 2)



The largest SMR units can produce close to 7 Msm³/d or 1 GW H₂. Therefore, two parallel SMR trains would be sufficient. Two parallel WGS and CO₂ removal trains will also be needed.

The main advantages/disadvantages are summarised in [Table 3.9](#).

OPTION	MAIN ADVANTAGES	MAIN DISADVANTAGES
1 SMR LP CC	Self-sufficient with power	Lower efficiency
2 SMR HP+LP CC	High efficiency. Electric power can also be produced onsite without changing the overall process performance.	Two different CO ₂ capture plants needed, which adds complexity and cost

Table 3.9: Comparison of SMR concepts. Main advantages and disadvantage

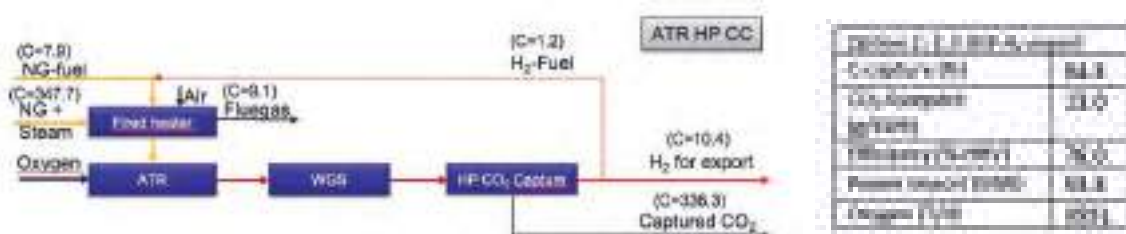
Increased capture rate above 91% in option one is unlikely since all CO₂ is captured at low pressure and it is very close to what is theoretically possible in such a plant. In option 2, only one third of the CO₂ is captured at low pressure. In this case, a higher capture rate can be obtained using some hydrogen as fuel in the reformer. However, this will cause a lower efficiency.



CO₂ capture from an ATR based hydrogen process – 1.5GW H₂ production

As with SMRs there are multiple alternatives for ATRs which can achieve high capture rates. The most promising concepts are presented diagrammatically in Figure 3.29 with a brief description of each alternative.²⁶

ATR OPTION 1: OXYGEN FIRED ATR WITH SYNGAS HIGH PRESSURE CO₂ CAPTURE (ATR HP CC)



ATR OPTION 2: Oxygen fired ATR with Gas Heated Reformer (GHR) and Syngas High Pressure CO₂ capture (ATR+GHR HP CC)

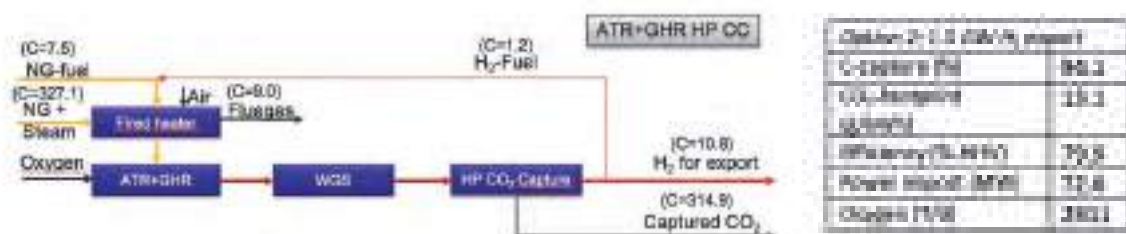


Figure 3.29: Schematics of the two ATR concepts including the carbon mass balance (CO₂ losses in the condensate around 1t/h)

²⁶ N.B. C-Capture (%) / Efficiency from UniSim Output, CO₂ Footprint based on NG parameters in Table 3.10. Power consumption (100%) included in efficiency numbers.

ATR OPTION 1: Oxygen fired ATR with syngas high pressure CO₂ capture (ATR HP CC)

This is a traditional ATR based process where mainly all heat is being recovered by generating steam.

All CO₂ can be generated at high pressure making it more efficient to capture than if it is released at low pressure in a steam reformer flue gas. The ATR is significantly more compact than a traditional SMR, but a large oxygen plant will be needed. Single train configuration up to 1.5GW H₂ is possible.

The high temperature in the ATR reactor is an advantage as most methane is converted to CO and H₂. This allows for a high CO₂ capture rate where the majority of CO can be converted to CO₂ in the water gas shift section. The high temperature also allows for operating the reformer at high pressure. This is an advantage if hydrogen must be delivered at a high pressure (as required for the H21 NoE concept).

ATR OPTION 2: Oxygen fired ATR with GHR and Syngas high pressure CO₂ capture (ATR+GHR HP CC)

To further increase the efficiency of the process and at the same time keep a high carbon capture rate, it is recommended to include a gas heated reformer in parallel with the ATR reactor to make better use of the heat in the hot syngas. In this case the oxygen plant size will be reduced with about 18% saving on both CAPEX and energy.

An optimised process concept using this technology can capture about 94% of carbon used in the production of hydrogen with an overall efficiency of about 80% (if all power needed is imported as renewable power). Due to the heat exchanger reformer being operated at a lower temperature than the ATR, less methane will be converted to hydrogen. This explains the somewhat lower CO₂ capture rate in this case compared to ATR option 1.

A better alternative would be to include the GHR in series with the ATR to keep the high methane conversion as in option 1. This technology, however, is less mature (demonstrated in a scale 10% of that needed here) and not assumed in this study. Should be evaluated in future studies.



Performance of the different ATR options are summarised in [Table 3.10](#) and main advantages and disadvantages are summarised in [Table 3.11](#).

1.5GW H ₂ EXPORT	OPTION 1 ATR	OPTION 2 ATR+GHR
CO ₂ capture (%)	94.5	94.1
CO ₂ footprint (g CO ₂ /kWh)	13.0	13.1
Efficiency (%-HHV)	76.0	79.9
NG consumption (MW)	1,919	1,805
Electric Power Import (MW)	53.8	72.6
Steam production (t/h)	683	527
CO ₂ export (t/h)	336	315
H ₂ purity (%)	98.4	98.4
H ₂ export pressure (bar)	80	80
Oxygen (t/d)	3,551	2,911

Table 3.10: Comparison of ATR alternatives

Installing a gas heated reformer significantly improves the efficiency and keeps a low carbon footprint. It will also reduce the consumption of oxygen and therefore the costs associated with O₂ production.

The area needed for a 1.5GW facility is approximately 15-20 ha. An example is the Oryx GTL plant in Qatar where the ASU and ATR part of the plant equals 4GW of hydrogen production, ([see Figure 3.31](#)).





Figure 3.30: Oryx GTL plant in Qatar

To present a credible option for H21 NoE where the preferred fuel is hydrogen for all energy supply, Option 2 has been selected (but a final selection can only be determined as part of a detailed future FEED study).

OPTION	MAIN ADVANTAGES	MAIN DISADVANTAGES
1 ATR HP CC	Low C-footprint	Lower efficiency High oxygen consumption
2 ATR+GHR HP CC	Low C-footprint and higher energy efficiency	High power import/power consumption

Table 3.11: Comparison of ATR concepts. Main advantages and disadvantages



3.4.4 Hydrogen process selection

Energy efficiency and CO₂ capture rate is the main driver for selecting preferred technology and therefore SMR option 2 is considered superior to SMR option 1. The ATR in combination with a gas heated reformer yields some improved energy efficiency and cost reduction. As such, ATR option 2 is selected for further evaluation.

The next step in the evaluation was to establish cost estimates for the preferred ATR and SMR schemes.

The hydrogen process used in the ATR cases have been developed and cost estimated by UK based KBR (a world leading technology supplier and EPC for the ammonia, hydrogen and oil/gas market) for delivering 10 million sm³/hydrogen per day at 45 bar, equal to about 1.46 GW (HHV). The plant is scaled up to 1.5 GW and equipment for compression of hydrogen from 45 bar to 80 bar are included.

Equinor internal cost estimation methodology has been aligned with the KBR estimate to develop costs for other hydrogen process options and hydrogen production capacities. The equipment lists provided by KBR are scaled and equipment cost is estimated using Aspen In-Plant Cost Estimator V9.

This is a well-documented cost estimation tool being used by many engineering companies worldwide.

The KBR reference plant estimate cost basis is 2nd quarter 2017 but adjusted to 2018. The overall accuracy is +/-40%. Based on UniSim modelling, detailed flowsheets have been developed and equipment lists are scaled so the plants can export 1.5 GW hydrogen.

In the SMR case, the ATR, GHR, fired heater and ASU units are removed and replaced with a traditional steam reformer. The other systems like CO-shift, syngas CO₂ removal, CO₂ compression, steam and power generation systems are scaled to required sizes.

The ATR cost estimate is more accurate than the SMR estimates since it is mainly scaled up from the KBR estimate. It is used in the assessment of the different large-scale hydrogen production scenarios.

A summary of CAPEX is presented in [Table 3.12](#) for the ATR and SMR case.

1.5GW H ₂ EXPORT (£M)	ATR+GHR HP CC	SMR HP+LP CC
OPTION	ATR 2	SMR 2
Equipment	197	206
Bulk	140	183
Indirects	38	41
Construction	236	260
Home office	16	22
CMT	23	24
Other	8	9
EPC (£M)	658	745
Owners cost	80	93
Equinor project management	33	40
Insurances	18	26
Contingency	158	178
TOTAL PROJECT COST	947	1,082

Table 3.12: Cost comparison of selected ATR and SMR hydrogen production technologies

Owners cost includes: Additional site work, substation and tie-in with HTrS and NTS (see Section 5.0).

The following owner's costs are excluded in the EPC cost estimates:

- Contractor fee and profit
- Operation and Maintenance Spares
- Licensor Basic Engineering Fees and License/Royalty Fees
- Insurances, bonds, guarantees (including Contractors All Risk, marine and module transport)
- Demolition, removal and/or disposal of hazardous materials (asbestos, lead, soils)
- Development of any additional infrastructure to support construction and operation
- Costs due to plant evacuations and work interruptions
- Currency risk
- All import/export duties
- Funded liabilities
- Land costs
- Sales tax
- Government permits (state and federal)
- Forward escalation

The best ATR and SMR concepts are not very different in efficiency. However, due to higher conversion of methane in an ATR reactor than in a SMR reactor (due to significant higher reforming temperature), it is easier to achieve higher carbon capture rates and a lower CO₂ footprint.

For a facility with many hydrogen production trains (10GW or more), it is recommended to have a centralised power plant to reduce complexity and costs. The preferred option is a centralised hydrogen fired power plant. Another option is to build a natural gas fired power plant with flue gas CO₂ capture.

The main benefits of the ATR over the SMR are summarised below:

- ATR can be operated at a higher temperature and pressure. This allows high conversion of methane and high syngas pressure at a lower steam to carbon ratio. High syngas pressure is an advantage if the hydrogen produced is needed at high pressure to reduce the need for hydrogen compression as is the case with the H21 NoE concept. In addition, this allows smaller physical unit dimensions but higher energy outputs
- Close to 95% of all CO₂ can be generated at high pressure which makes it more cost efficient to capture than if it is being released at low pressure in the SMR flue gas
- The study shows that the CO₂ footprint for the ATR based process is about 13 g CO₂/kWh compared to about 20 g CO₂/kWh for SMR, which is a considerable improvement.
- ATR based design instead of SMR would results in reduced safety risks by having fewer numbers of vulnerable components for leaks/ failure and lower operational complexity
- An ATR based process technology also largely improves on the hydrogen plant's ability to meet quicker start up and shutdown

CAPEX for the selected ATR alternative is around 12-13% below the best SMR alternative but will need 37 MW more electric power. ATR based concepts may have some better cost reduction potential than SMR based concepts if scaling up to larger capacities since ASU's can be scaled up to serve around 2.5 GW H₂ production. Comparison of the best ATR and SMR options are summarised in [Figure 3.13](#).

1.5 GW H ₂ PRODUCTION	ATR OPTION 2	SMR OPTION 2
Carbon capture rate (%)	94.1	91.2
CO ₂ footprint (g CO ₂ /kWh)	13.1	20.5
Efficiency % (HHV)	79.9	79.5
CAPEX (£m) – Total	947	1,082
Electric Power Import (MW)	72.6	35.6
CAPEX £/kWh _{2HHV})	631	721
Area (ha)	15-20	35-40
Configuration	1 ATR train + ASU	2 SMR trains

Table 3.13: Comparison of the best ATR and SMR concepts

3.4.5 Supply chain maturity

The hydrogen reforming technology market is mature and highly competitive. It comprises of 8-10 international technology providers (see Table 3.14) that deliver process design packages, special equipment, catalysts and, in some cases, complete turn-key project delivery.

COMPANY	TECHNOLOGY EXPERIENCE			INDUSTRY SEGMENT EXPERIENCE		
	SMR	ATR	GHR	AMMONIA	METHANOL	HYDROGEN
Air Liquide	•	•			•	•
Johnson Matthey	•	•	•	•	•	•
Kelloggs Brown & Root	•	•	•	•		•
Linde	•	•				•
Technip	•					•
Thyssenkrup	•	•		•	•	•
Haldor Topsoe	•	•	•	•	•	•
Amec Foster Wheeler	•					•

Table 3.14: Providers of natural gas-based hydrogen technology

In addition to the above companies, there are numerous EPC contractors who have experience in the engineering and construction of hydrogen facilities.

For the H21 NoE Hydrogen Production Facility, the ATR option 2 with GHR has been selected as hydrogen production technology for design and commissioning of the 10-15GW mega scale hydrogen facility.

In the FEED phase ATR and SMR based hydrogen production concepts will be further analysed in addition to supply of power.



3.5

Gasification

As described in [Section 3.1 'History of H₂ Production'](#), gasification of oil and coal is the other main global method of large-scale hydrogen production.

In the UK, coal as a feedstock has now been almost totally phased out and therefore considered impractical as a credible feedstock for H21 NoE. Additionally, the logistics and economics associated with such a process make it unattractive in the UK.

However, for completeness and relevance for other countries with low availability of natural gas but access to significant CCS, a short section on coal gasification has been provided.

3.5.1 Gasification references

Large scale (>500 tonnes/day of coal per train equivalent to 100 MW of hydrogen production capacity) coal gasifiers have been in operation since the 1930s. Notable examples are the SASOL gasification complexes in South Africa (originally built in the 1950s and 1960s and still operating over 80 Lurgi fixed bed dry ash gasifiers today) and the Great Plains Syngas Plant in the USA which has operated with high availability since the 1980s. Over the last 70-80 years gasification technologies have been developed, deployed, refined and subsequently rendered obsolete by newer processes.

The Gasification & Syngas Technologies Council maintains a database of commercial gasification plants and the 2015 update identified over 700 units operating worldwide (2). The main technologies in operation were:

- GE (formerly Texaco) – 180 units
- Shell – 110 units
- Lurgi (now owned by Air Liquide) – 130 units (+ over 50 of the similar SEDIN design)
- ECUST (East China University of Science & technology) – 25 units
- Other design (Siemens, E-Gas, Prenflo, BGL, Choren, MCSI, SES and others) – 30-40 units



The majority of these gasifiers are in China where the availability of low-cost coal and the lack of competition from other fuels has favoured their deployment. Construction of large scale entrained flow gasifiers (Texaco and Shell designs) commenced in China in the early 2000s, hence a number of these plants have now been in operation for over 10 years.

Commercially installed coal gasification technologies typically exceed 100 MW_{th} in train size. Fixed bed gasifiers such as the Lurgi and BGL designs exist up to approximately 250 MW_{th}, whilst entrained range up to 500 MW_{th} (Shell) and 850 MW_{th} (Siemens, who are developing a 1,000 MW_{th} design).

For a baseload plant, large unit sizes may offer an advantage. However, it should be noted that the turndown of many gasifier designs is limited and that the added reliability and easier maintenance offered by multiple smaller gasifiers may be preferred. A good example is the Great Plains Syngas plant where 14 Lurgi dry ash gasifiers (each unit with a capacity of 150-200 MW_{th}) are installed and which has operated at high availability throughout its lifetime. This approach can add to costs but offers advantages by eliminating scale-up uncertainties, standardisation and availability. This last feature is a major benefit as the high temperature and aggressive chemistry associated with gasification processes means that routine maintenance and refurbishment is to be expected and spare gasifier modules will usually be specified. For most gasifier designs this modularity extends to the initial gas clean-up and cooling of the raw gas stream.

3.5.2 Description of the technology

Gasification is conversion of a feedstock, usually coal or heavy oil, to synthesis gas (syngas) and follows the same principles as the natural gas-based ATR technology. In greatly simplified terms, heat generated from combustion of a fraction of the feedstock is used to drive endothermic gasification reactions to generate the main syngas components – Carbon Monoxide (CO) and H_2 . CO is converted into hydrogen in the WGS and CO_2 removed at high pressure by a solvent.

The biggest difference between a coal gasification process and an ATR based process is the handling of a solid feed to the reforming process (gasifier), amount of CO_2 to be captured, oxygen consumed and the need for handling a highly H_2S rich syngas. Hydrogen production by gasification of coal presents the general principles of hydrogen production from coal gasification.

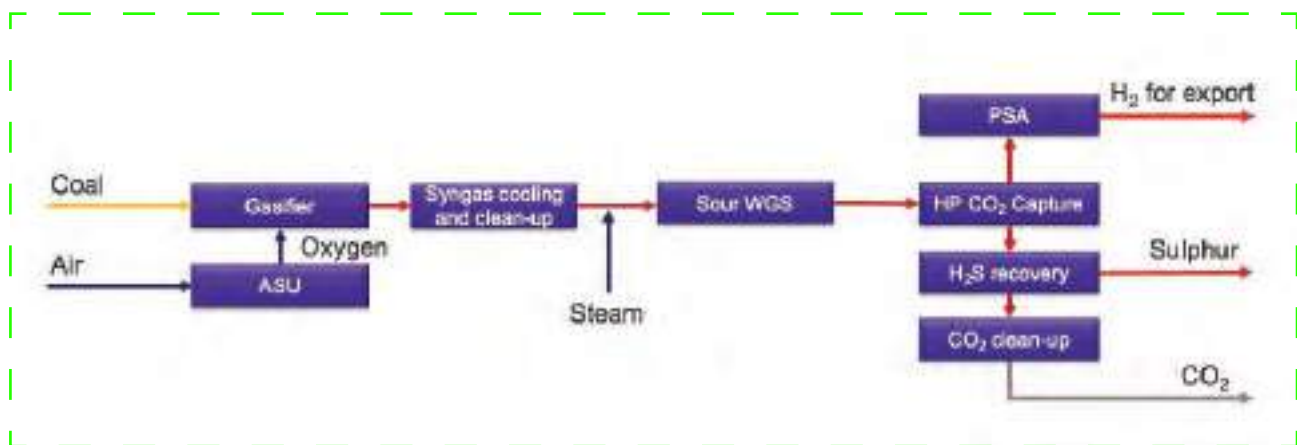


Figure 3.31: Hydrogen production by gasification of coal



Main gasifier types

Large scale commercial coal gasifiers can be broadly divided into three types; entrained flow, fluidised bed or moving bed.

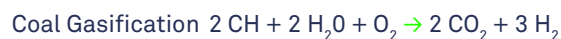
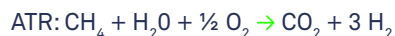
Entrained flow operate at high temperatures, typically between 1,400 and 1,800°C. Finely ground coal is continuously fed, either as a water slurry or using a carrier gas, into the gasifier via a 'burner' where it combines with the gasifying agents (air, oxygen and/or steam) and rapidly reacts.

Moving bed (sometimes referred to as fixed bed) is the longest established means of converting coal to gas at temperatures up to 1,000°C. The gasifier contains a bed of lump coal which is gradually consumed by the gasification reactions. The gasifying agents are typically fed to the bottom of the coal bed. Gasification agents like fresh coal is locked into the gasifier and mechanically distributed on the top of the coal bed.

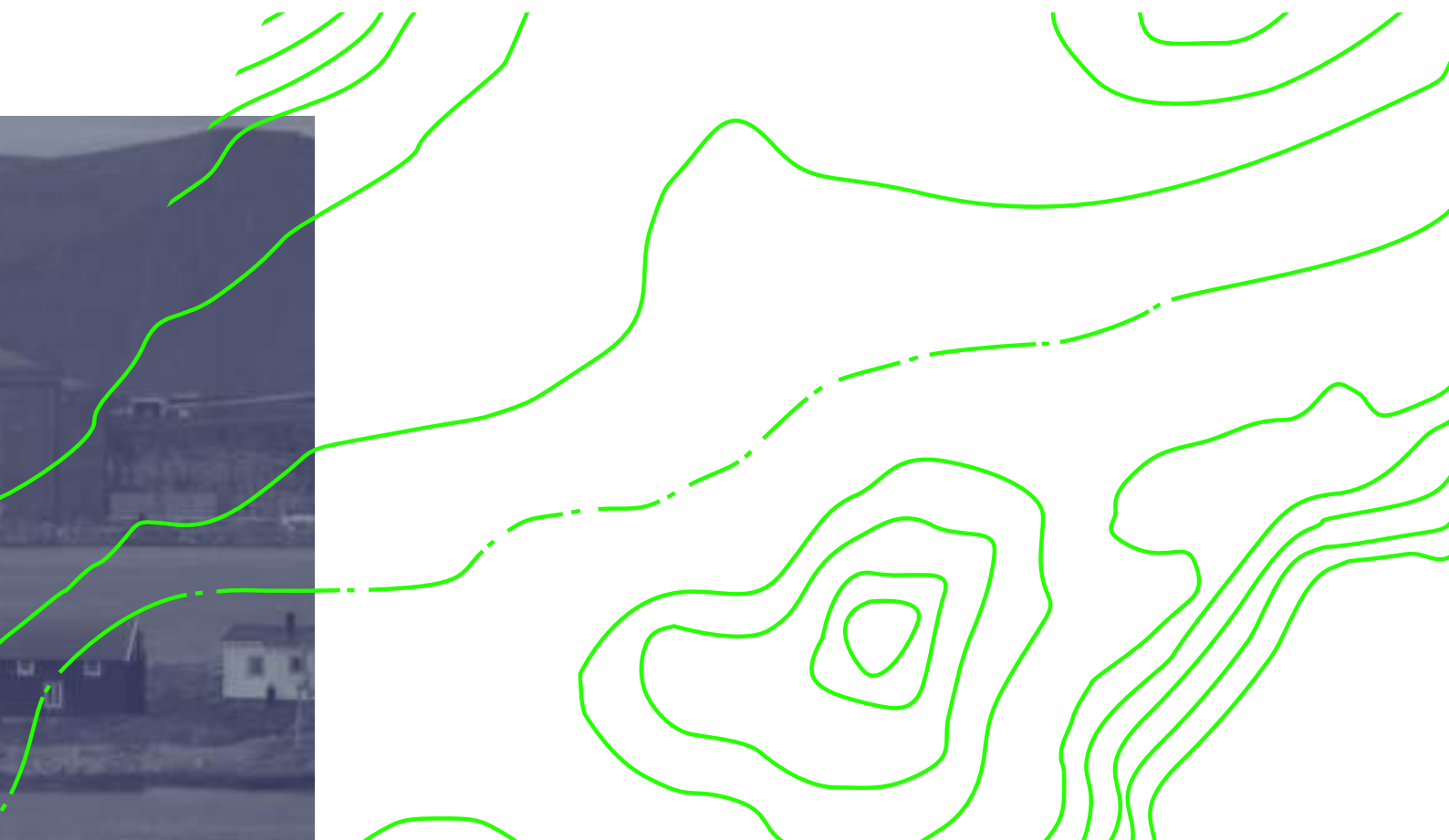
Fluidised bed conversion of the coal takes place in a fluidised bed at temperatures of between 1,000 and 1,500°C. The gasifying coal is contained as a fluidised bed maintained by the flow of gasifying agents and product gas.

As the largest gasifiers in operation are in the range of 250 and 300 MW hydrogen (equal to 500 MW_{coalLHV}), a hydrogen facility of 1.5 GW would need 5 to 6 gasifier units operating in parallel. This provides flexibility but also increased cost.

Comparing the overall conversion reaction of coal and methane, twice as much oxygen and steam is needed with twice as much CO₂ needed to be captured, transported and permanently stored.



This results in significant increase of capital and operational costs compared to a natural gas-based ATR scheme. **The main motivation for coal-based solution is the lower cost of coal compared to natural gas and lack of gas availability.**



The CO₂ capture rate potential depends heavily on gasification type. An analysis of gas composition after the high-pressure CO₂ removal shows that only entrained flow type gasifier can meet a CO₂ capture rate of 90% (Table 3.15).

	SHIFTED GAS COMPOSITION WITH 99% CO ₂ REMOVED (NORMALISED)						MAX % CARBON CAPTURE
	H ₂	CH ₄	CO	CO ₂	N ₂ AND AR	C2+	
MOVING BED SLAGGING	82.4	7.4	5.9	0.5	2.9	0.8	77.9
MOVING BED DRY ASH	80.6	12.3	2.6	0.7	2.3	1.4	78.7
ENTRAINED FLOW, SLURRY FEED, SINGLE STAGE	91.8	0.1	5.2	0.6	2.3	0.0	91.2
ENTRAINED FLOW, SLURRY FEED, TWO STAGE	90.5	0.4	5.4	0.6	3.2	0.0	90.0
ENTRAINED FLOW, DRY FEED	88.6	0.0	6.3	0.6	4.5	0.0	89.2
FLUIDISED BED	87.8	4.7	5.2	0.7	0.6	1.0	84.1

Table 3.15: CO₂ capture rate of coal gasifiers

Gas clean-up and cooling

The initial downstream gas clean-up (tar, oil and dust removal) and cooling is specific to individual gasification technologies. The lower temperature syngas from fixed bed gasifiers requires immediate water washes and quenching to protect downstream process equipment from the high levels of tar, oil and dust. The high temperature syngas exiting an entrained flow gasifier is free of tar and oil but will carry dust and molten slag droplets. Radiant coolers which maximise high grade steam recovery are possible on most entrained flow technologies but are large, expensive and have suffered from damage by slag impingement or metallurgical issues. As a result, many entrained flow gasification plants opt for a lower cost quench cooling system or a combination of partial quench downstream of a smaller radiant cooler.

Sour water gas shift

For most downstream applications, the ratio of H₂ to CO in raw syngas from the gasifiers must be adjusted using the shift process in a series of catalyst beds. This technology has been in use for over 50 years and a range of catalysts (low temperature, high temperature and sour) are manufactured and offered with guarantees by manufacturers such as Johnson Matthey, Clariant (formerly SudChemie) and Haldor-Topsoe.

High pressure CO₂ removal

Physical solvent processes such as Benfield, Rectisol and Selexol absorb acid gas components (including CO₂) in the liquid solvent as a function of their partial pressure. These processes are best suited for treating high pressure syngas streams. The solvents are regenerated by pressure reduction, temperature increase or both. Physical solvent plants have been in operation for over 50 years and there are more than 100 plants worldwide. They are well proven and very effective but their complexity may make them unsuitable for a small-scale plant.

At the Great Plains Syngas Plant, CO₂ is separated from the syngas in a Rectisol plant. Although the CO₂ separated at Great Plains is sour (i.e. it contains separated H₂S and other sulphur compounds), it is exported at high pressure for Enhanced Oil Recovery (EOR) at the Weyburn oilfield. CO₂ recovery from shifted syngas is also practiced at several chemical complexes in China to provide a feedstock for urea production.





Final clean-up

Since the gas leaving the shift section and CO₂ removal contains more than 5% CO, a final clean-up step is needed. The preferred technology is Pressure Swing Adsorption (PSA) as previously described in [Section 3.4](#).

By-products

In the CO₂ removal system H₂S is removed, however H₂S is toxic and must be further processed to sulphur for safe disposal. This is normally achieved in the Claus process which is a well proven technology in operation for more than 50 years.

In addition to sulphur, coal contains many elements, e.g. heavy/toxic metals, will be released during the process and end up as residue in the condensate and must be handled and disposed of safely. Further large amount of ash/slag will be generated as a by-product which can be used in the cement industry.

3.5.3 1.5GW facility based on gasification

Based on the design requirement for ATR and SMR based hydrogen technology a 1.5GW coal gasification-based hydrogen facility was developed with support from DNV GL. GE designed gasification technology was used in the analysis using Bituminous coal ($8.4 \text{ kWh/kg}_{\text{HHV}}$).

Key design parameters are given in below.

1.5GW H ₂ EXPORT	ENTRAINED FLOW GE GASIFICATION
Coal consumption (t/hr)	322.8
Coal consumption (MW_{HHV})	2,711
Efficiency (%-HHV)	55.3
CO ₂ footprint (g/kWh)	60.3
CO ₂ capture rate (%)	88.0
Electric power consumption (MW)	220-250
CO ₂ export (t/h)	663.5
H ₂ purity (%)	99.9
H ₂ export pressure (bar)	80

Table 3.16: Key design performance metrics for coal gasification

A cost estimate was established based on the same methodology and principles as the cost estimate for the ATR and SMR case.

The 1.5GW gasification facility comprises the following main features

- Two air separation unit of 3,500 t/d
- Six entrained flow GE gasifiers of $452 \text{ MW}_{\text{CoalHHV}}$
- Two syngas cooling, sour gas and HP CO₂ removal units
- One common Claus plant for H₂S recovery
- A steam turbine power generation unit of 240 MW
- Hydrogen and CO₂ compression
- Pits for waste water, sulphur, ash/slag
- Chemical storage
- Substation for power import
- Coal receiving by train and storage
- The 1.5GW gasification-based hydrogen facility was estimated to £2.5bn equal $\text{£}1,691/\text{kW}_{\text{H}_2\text{HHV}}$ with a breakdown given in the following table

1.5GW H ₂ EXPORT (£M)	ENTRAINED FLOW GASIFIER
Equipment	514
Bulk	387
Indirects	102
Construction	658
Home office	47
CMT	63
Other	23
EPC	1,794
Owners cost	200
Project management	90
Insurances	54
Contingency	399
TOTAL PROJECT COSTS	2,537

Table 3.17: Capital cost breakdown of a 1.5GW coal gasification hydrogen facility

With reference to the Great Plains gasification plant of 2,000 MW_{th} – see [Figure 3.32](#) – the estimated total area for a 1.5 GW hydrogen facility is approximately 60–100 ha (0.6–1.0 km²).



Figure 3.32: Aerial photo of Dakota Gasification Company facility in the North Dakota US – equal 2,000MW



3.5.4 Gasification summary

Coal gasification and the subsequent processing of the syngas produced is an established technology and there are several hundred large scale installations operating in China. Large gasification plants also operate in South Africa (SASOL) and in North America. Several major vendors offer commercially proven gasification and gas processing technologies suitable for large scale H_2 production plants.

A coal to H_2 plant of 1.5 GW at a UK location is estimated to cost in the region of £2,537m or £1,691/ kW_{H_2HHV}

CO_2 emissions for the processes considered is estimated to 60 g/ kW_{H_2HHV} . Other combinations of coal types and gasification might achieve lower values.

An area of 60-100 ha is needed for 1.5 GW facility or 40-70 ha/GW.

Handling waste materials such as waste water, ash/slag and sulphur will be required.

An infrastructure for supply of coal is needed.

Building coal gasification plants at existing coal power plant sites can contribute to cost savings of 20-25%.

3.6

Ammonia

Ammonia is a hydrogen carrier with excellent properties for the transport and storage of hydrogen. Ammonia is liquid at -33°C at atmospheric pressure which is similar to LPG. Combined with a high hydrogen content on a mass and volume basis this makes storing of ammonia both feasible and cost effective.

Ammonia properties are shown in [Table 3.18](#).

PROPERTIES	VALUE	UNIT
H ₂ weight fraction	17.65	Wt.%
Boiling point @ 1atm	-33.5	°C
Liquid density (1atm/-33°C)	682	kg/m³
Molecular weight	17.03	g/mol
Autoignition temperature	630	°C
Boiling point @ 40bar	78.9	°C
Boiling point @ 20bar	51.5	°C
Energy density (HHV)	6.25	kWh/kg

Technology for ammonia production is well-established and will no doubt play an important part in a hydrogen economy in the longer term. Applications may include international trading of hydrogen, i.e. shipping (like the liquid natural gas market of today), storage alongside or instead of salt caverns and hydrogen production for areas/countries with limited alternative feedstock.

Ammonia was considered for H21 NoE for the following reasons:

- As an alternative to inter-seasonal storage ([see Section 3.7](#))
- As a hydrogen import source (like the role of LNG today) from renewables and other clean sources
- As strategically located hydrogen production alternative which would alleviate the need for some elements of storage

Table 3.18: Physical properties of ammonia



3.6.1 Ammonia production

Ammonia is produced through the Haber-Bosch process ($3\text{H}_2 + \text{N}_2 \rightarrow 2\text{NH}_3$) in large chemical plants. This is undertaken following production of gaseous hydrogen, for example produced via a steam methane/autothermal reformer, then catalytically reacting this hydrogen with nitrogen in the ammonia synthesis loop. Ammonia can be stored in large storage tanks (60,000 tonnes, 375,000 MWh) or shipped around the world in very large tankers (50,000 tonnes, circa 310,000 MWh). If the ammonia is being used as a storage medium or imported for future hydrogen production, it will subsequently need to be 'cracked' to regenerate the hydrogen.

The exothermic reaction to produce NH_3 from nitrogen (N_2) and H_2 (hydrogen) is undertaken as follows:



Low temperatures and high pressures promote a high degree of conversion to NH_3 but at low temperature the reaction proceeds too slowly to be acceptable at industrial scale. Optimal reactor temperatures

for the Haber-Bosch process are 350–500°C. There is a trade-off between cost of compression and reactor wall thickness, giving an optimum pressure in the range of 90 to 200 bar. Up to 20% conversion per pass is possible at optimal conditions. This means that a significant amount of gas must be recycled in the process depending on the content of inert gases. Traditionally an iron-based catalyst is used with reactor pressure around 140–200 bar.

The ammonia synthesis is presented schematically in Figure 3.33 and consists of the following process steps.

- Compression of make-up gas (N_2/H_2 mixture) and recompression of recycled gas.
- Reactor (converter) for generation of ammonia
- A separator system where ammonia is being condensed and separated from the gas phase at high pressure
- A refrigeration system where ammonia is cooled to -34°C and depressurised to 1 atmosphere

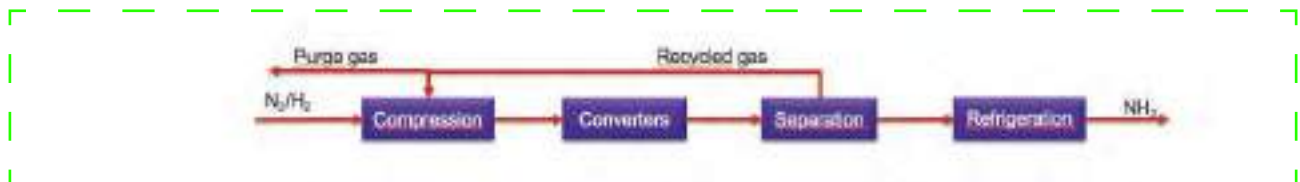


Figure 3.33: Simplified ammonia synthesis loop

The reaction between nitrogen and hydrogen to ammonia is exothermic, increasing the temperature to about 480°C in the first reactor. The gas is then cooled to about 360°C before entering a new catalytic reactor. Steam is being generated between the reactor steps and after the last reactor step. Steam is used in steam turbines to drive the different compressors.

The partially converted gas is cooled down in several heat exchangers to between °C5 and °C18 depending on the pressure. These include producing steam, heating boiler feed water (BFW), pre-heating converter feed streams, by cooling water and finally by ammonia chillers. Most of the generated ammonia is condensed and separated. Gas from the separator is re-compressed and mixed with compressed make-up syngas.

Separated NH₃ liquid pressure is then reduced in two steps. Firstly, down to 20-25 bar where dissolved gases are separated out and secondly, further down to ambient pressure where the temperature is reduced to 33.5°C. Liquid ammonia is then pumped to the storage vessel/tank/tanker. A minor part of liquid ammonia is evaporated in a let-down process (part of the refrigeration section) and is recompressed for recycling to the process.

The make-up gas to the synthesis loop contains some inerts mainly composed of methane and argon in conventional designs. To avoid accumulation of these components in the loop, a small purge stream is taken from the cooling train. This occurs at a temperature where the ammonia content in the gas is low. The purge gas is used as fuel in the steam reformer or fired heaters after the remaining ammonia is recovered from it.

For ammonia production from natural gas with CO₂ capture, the preferred scheme is an oxygen blown ATR like the concept for hydrogen production. The advantage is that the Air Separation Unit (ASU) produces nitrogen as a by-product which is needed in the ammonia synthesis. Since the hydrogen gas leaving the CO₂ removal contains traces of CO₂ and CO these must be converted into methane to avoid poisoning of the ammonia synthesis catalyst. This is achieved via reaction called methanation:



Methanation is standard technology used in ammonia production globally. The overall concept is shown in Figure 3.34.

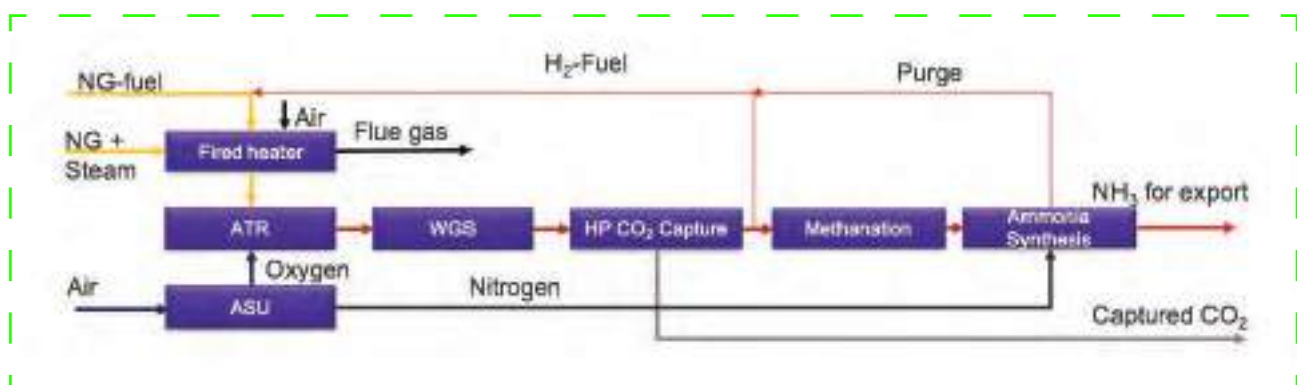


Figure 3.34: Ammonia production based on methanation



Ammonia based on renewables

Ammonia can also be produced from renewable sources via electrolysis ('green ammonia'). Low cost storage and transport of ammonia offers an opportunity to utilise availability of cheap renewable sources, e.g. solar power in hot climates. The advantage of solar energy as a renewable source are as follows:

- Solar produces Direct Current (DC) power which is the requirement for electrolysis (No AC-DC converter is necessary)
- Solar in hot climates provides stable and continuous power source compared to wind which is, by its nature, an intermittent power source. Obviously, there is a power production swing during the day, i.e. between daylight and hours of darkness at night. This is where Li-Ion batteries could act as a buffer between the solar power plant and the electrolyser to level the power production. This would allow for the electrolyser to operate continuously and make hydrogen/ammonia production more economic.

A renewable scheme based on electrolysis will require installation of an air separation unit for producing the required nitrogen for the ammonia synthesis. However, since the hydrogen from an electrolyser is ultra pure, some simplification of the ammonia synthesis loop can be achieved. The overall concept for producing ammonia via electrolysis is shown in [Figure 3.35](#).

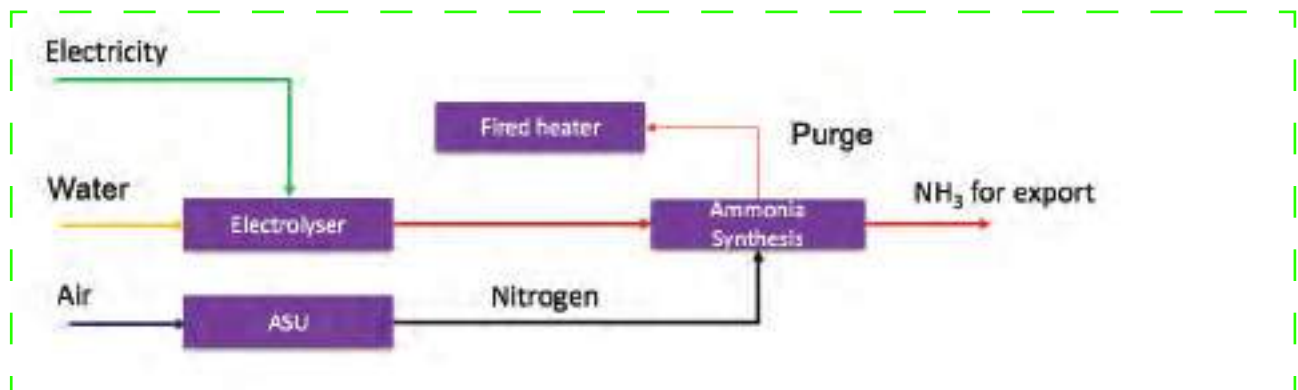


Figure 3.35: Ammonia by electrolysis



Ammonia based on gasification

Ammonia can be produced as part of coal gasification; this is the specific option being considered by the Carbonnet Project run by the Victoria Government in Australia. This may represent a credible source of internationally traded 'clean' hydrogen in the longer term for the UK. However, it is not considered practical in the context of the H21 NoE project as much more efficient, cheaper and cleaner systems are available directly to the UK via natural gas.

The gasification-based ammonia production scheme can either be designed by adding the ammonia synthesis loop to the hydrogen production scheme or, by cleaning the syngas, using a nitrogen wash. The first concept is shown in [Figure 3.36](#).

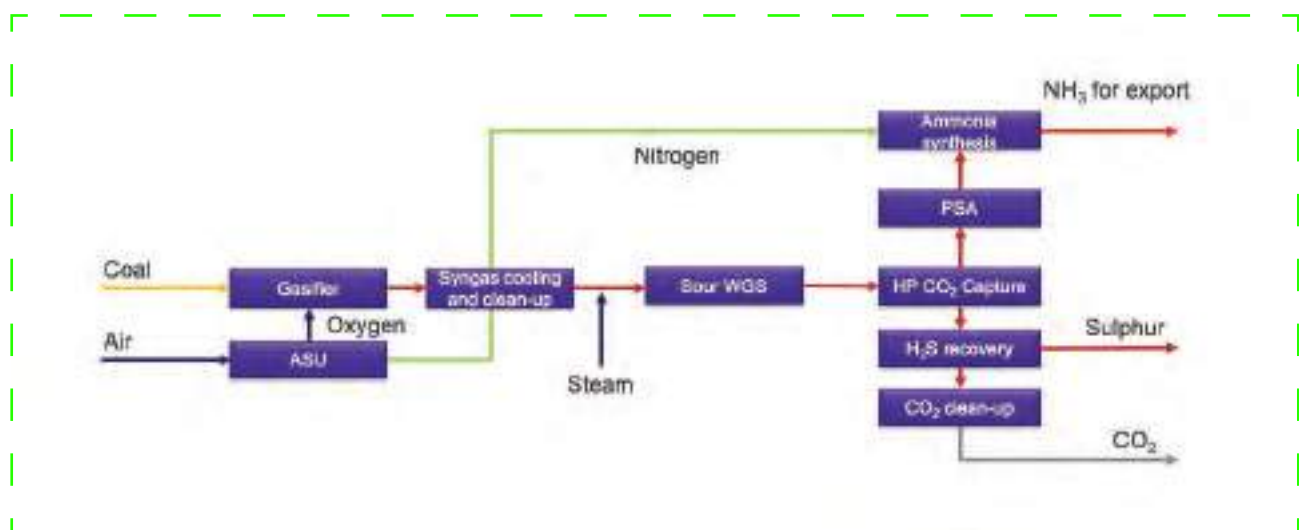


Figure 3.36: Ammonia by coal gasification

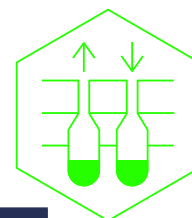
Supply chain considerations

There are several technology vendors for the ammonia synthesis loop such as KBR, Topsoe and Uhde. Variations are related to factors such as reactor design, catalyst and operating pressure. Two of these are summarised briefly below:

- KBR's KAAPplus™ process. The development of cost-efficient ruthenium catalysts has been a driver to reduce the pressure to 90 bar in a single compressor casing and the synthesis loop purge is mixed with the make-up gas upstream of the purifier
- Uhde dual-pressure process developed by Thyssenkrupp and Johnson Matthey. The first plant to apply this process was the SAFCO IV ammonia plant in Al-Juail, Saudi Arabia with a capacity of 3,300t/d started up in 2006. In this process compression of make-up gas is carried out in two steps. Due to the cold NH_3 separation (below -10°C), the HP compressor operates at a much lower temperature than normal allowing efficient compression up to 210 bar increasing the conversion to ammonia.

Key performance

An ammonia synthesis design for the ATR, electrolyser and gasification hydrogen scheme was developed by Equinor. This was modelled and key performance data and design data for cost estimation was developed.



5,000 T/D NH ₃ EXPORT 1.3 GW NH ₃	ATR+GHR BASED	ELECTROLYSER BASED	COAL GASIFICATION
CO capture (%)	92.4	N/A	87.5
CO ₂ footprint (g/kWh NH ₃)	20.6	0	65.1
Energy consumption (MW _{HHV})	1,831	2,090	2,642
Efficiency (%-HHV)	71	62.2	49.2
CO ₂ export (t/h)	298	N/A	612

Table 3.19: Performance figures – generic ammonia plant types



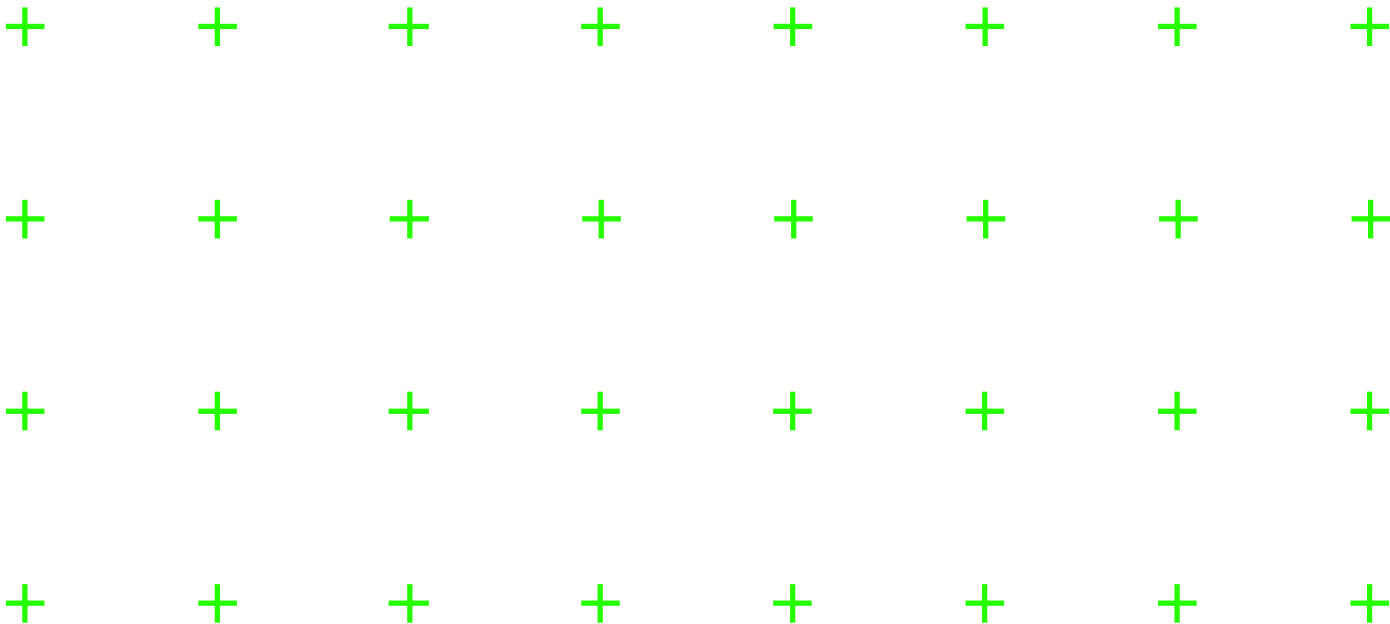
3.6.2 Capital cost

The ammonia cost estimate is based on an internal study using the appropriate hydrogen concept as presented in this section for electrolyzers, ATR and coal gasifications (i.e. each based on a 1.5 GW plants). These were integrated with a generic ammonia synthesis loop like the Topsøe synthesis loop. Owners cost includes: jetty, tank farm, loading facility, tie-in with natural gas system, substation and additional sitework.

Table 3.20 shows a capital cost breakdown of a 5,000 tpd (1.3 GW) Ammonia facility based on ATR, electrolysis and coal gasification technology.

5,000 T/D NH ₃ EXPORT 1.3GW NH ₃	ATR+GHR BASED	ELECTROLYSER BASED	COAL GASIFICATION
Equipment	295	1,065	592
Bulk	234	164	462
Indirects	60	40	119
Construction	384	296	776
Home office	27	26	56
CMT	33	31	71
Other	13	10	27
EPC (£M)	1,047	1,632	2,103
Owners cost	300	420	420
STL project managent	52	58	105
Insurances	31	53	63
Contingency	286	310	505
TOTAL PROJECT COST	1,717	2,473	3,196

Table 3.20: Capital cost figures – generic ammonia plant types



3.6.3 Ammonia cracking to hydrogen

The principles of ammonia cracking are based on a technology platform like steam methane reforming. Small scale units are in operation at low pressure. For large scale application the technology has been used predominantly in heavy water production (i.e. water in which the hydrogen in the molecules is partly or wholly replaced by the isotope deuterium) for use as a moderator in nuclear reactors. Arroyito Heavy Water Production Plant in Argentina is based on a method known as 'Monothermal Ammonia-Hydrogen Isotopic exchange'. In this plant more than 3,000 t/d of ammonia is cracked into a hydrogen-nitrogen mixture and corresponds to 0.7 GW of hydrogen.

For the Arroyito plant, Haldor Topsoe was selected as supplier (no other suppliers are known). However, in principle all suppliers of SMR technology can deliver a large-scale ammonia cracker facility.

Description of technology

The technology for converting ammonia into hydrogen and nitrogen molecules is called either ammonia cracking, ammonia dissociation or ammonia splitting. Ammonia cracking will be used in this report. The key advantage of ammonia cracking compared to other hydrogen processes is that it does not require any steam/water. Most of the heat can be used for evaporating and heating ammonia before cracking. This makes it possible to design a very flexible, robust and energy efficient concept.

Ammonia dissociates (cracks) into hydrogen and nitrogen according to the reaction:



The cracking rate depends on temperature, pressure and the catalysts being used. Ammonia cracking is an endothermic reaction, (i.e. it will consume a significant amount of heat). About 20% of produced hydrogen will be needed as fuel. Since cracked ammonia, (i.e. hydrogen) contains more energy than the ammonia feed, an efficiency above 86% (HHV) will still be possible (electric power consumption included).



A nickel, cobalt or ruthenium based catalyst can be used to crack ammonia. Including other components in the catalyst could increase conversion at lower temperature allowing reduced use of heat. The cracker process is quite simple but the need for efficient heat recovery and purification of cracked ammonia adds complication. Since very hot gases are being generated, heat recovery is an important part of the process. Steam is not needed in the production and it is essential that all available heat can be used in the preheating of the different process streams to minimise energy consumption.

For a UK gas grid conversion, it is likely that we will export produced hydrogen into the Hydrogen Transmission System (HTS), (see [Section 5.0](#)) at between 17 and 80 bar depending on the use. If cracked hydrogen must be exported at 80 bar a rather high cracker pressure would be optimal (around 40 bar) to reduce the need for hydrogen compression, but ammonia slippage will increase.

An efficient cracker process is illustrated in [Figure 3.37](#). The front-end process includes pumping of liquid ammonia to the required pressure, evaporation of ammonia and preheating. The temperature at which ammonia is being evaporated depends on the pressure. A low pressure allows use of ambient air as the heating medium. In a UK gas grid conversion, a higher cracker pressure is needed (20-40 barg) and ammonia needs to be evaporated above ambient temperature. Total conversion of ammonia is about 99.5% at 40 barg and 900°C.



Figure 3.37: Ammonia cracker process block-diagram

3.6.4 Capital cost

Ammonia cracker

Based on UniSim modelling, detailed flowsheets have been developed and equipment lists are scaled so the plants can export 1.5GW hydrogen. Equipment cost is estimated using Aspen In-Plant Cost Estimator V8.4. Installed cost is estimated using the same assumptions (factors) as used for the SMR based hydrogen plants. Owners cost includes: ammonia receiving facility, jetty, tank farm, substation and tie-in with HTS. A CAPEX summary is shown in [Table 3.21](#).

1.5GW H ₂ EXPORT	17 BARG H ₂ EXPORT	80 BARG H ₂ EXPORT
£M		
NH ₃ cracker case	1	2
Equipment	87	90
Bulk	69	72
Indirects	14	14
Construction	61	65
Home office	4	4
CMT	9	9
Other	3	3
EPC (£M)	245	258
Owners cost	30	32
Project managent	12	13
Insurances	7	7
Contingency	59	62
TOTAL PROJECT COST	353	372

Table 3.21: Cost estimate for the ammonia cracker unit bar

The cost estimates indicate that producing hydrogen by cracking ammonia will cost in CAPEX about one third of a hydrogen plant but the cost of NH₃ feed will be much higher than cost of natural gas feed. This is because the cost of the ammonia will include the original feedstock plus all the upstream ammonia production facilities and shipping requirements.

3.6.5 Ammonia summary

There are several options for large scale ammonia production and subsequent cracking. Each has different benefits which should be assessed as part of any future FEED study where ammonia is considered a critical component. An assessment of two ammonia cracking cases are provided in [Table 3.22](#).

1.5 GW H ₂	NH ₃ IN (MW _{HHV})	POWER IMPORT (MWEL)	EFFICIENCY ¹⁾ NH ₃ TO H ₂ (%-HHV)	EFFICIENCY ²⁾ CH ₄ TO H ₂ (%-HHV)	CO ₂ FOOTPRINT ³⁾ (G/KWH)	H ₂ EXPORT PRESSURE (BAR)
Case 1	1689	1.8	88.7	63.0	22.4	17
Case 2	1695	17.7	87.5	62.2	22.4	80

Table 3.22: Comparison of most efficient cracker concepts

- **1)** Efficiency includes cracked NH₃ used as fuel in cracker
- **2)** Total efficiency from natural gas via ammonia to hydrogen assuming 65% NH₃ plant efficiency (LHV)
- **3)** Assuming 92.4% carbon capture rate in ammonia plant

If, assuming 71% efficient (LHV) ammonia production, the overall efficiency will be about 62% (HHV) compared to ATR/SMR's at 79-80% (HHV) ([see Table 3.13 in Section 3.3](#)) assuming direct conversion from natural gas to hydrogen and hydrogen export pressure at 80 bar (case 2). The overall CO₂ footprint will be around 22 g/kWh. An increase from about 13 g/kWh for direct conversion based on ATR technology.



3.7

Summary of hydrogen production technologies

Hydrogen production from natural gas (ATR, SMR), electrolysis, coal gasification and ammonia cracking was analysed.

The key performance data and capital cost estimates are given in order of preference for the H21 NoE project in [Table 3.23](#).

	ATR	SMR	ELECTROLYSIS	GASIFICATION	NH ₃ CRACKING
Energy Efficiency (%-HHV)	79.9	79.5	70.9	55.3	87.5-88.9
CO ₂ footprint	13.1	20.5	0	60.3	0
Specific cost (£/kW _{HHVH2})	631	721	1,160	1,691	235-248
Area (ha/GW)	10-15	25-30	30-40	40-70	8-10

Table 3.23: Summary of key performance data for large scale hydrogen technologies

Based on the key performance data it is possible to establish hydrogen production curves using the economic pre-tax assumptions given in [Table 3.24](#).

ASSUMPTION	
Operational (hours per year)	8,500
Discount rate (%)	8
Project lifetime (years)	25
O&M cost	3% of CAPEX
CO ₂ T&S cost	£10/t

Table 3.24: Economic assumptions used for evaluating hydrogen production technologies



This can then be compared against the UK Governments Department of Business Energy and Industrial Strategy (BEIS) data for UK future coal, gas, electricity and CO₂ prices in 2,035 presented in table 3.25.

	LOW	CENTRAL	HIGH
NATURAL GAS (£/MWH)	13	23	28
COAL (£/MWH)	7	9	14
ELECTRICITY (£/MWH)	45	60	70
CO ₂ price (£/tonnes)	40	79	119

Table 3.25: Energy and CO₂ price prognosis from BEIS

In the figure below hydrogen production cost curves are presented including the cost range given by BEIS. In addition, a curve showing the impact for electrolysis assuming development succeeds in bringing costs down by 50% and improving efficiency by 10% (yellow line).

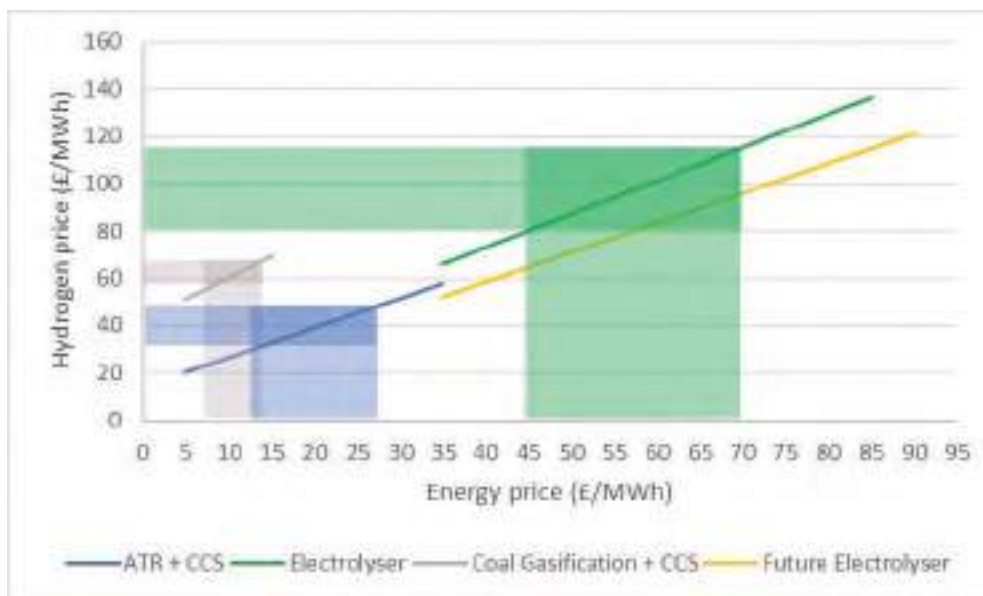


Figure 3.38: Cost curves for hydrogen production technologies

The green bar is BEIS future electricity prices, blue is future gas prices and grey is future coal prices.

With BEIS future prices as the basis it is clear that a natural gas hydrogen route offers the lowest cost to the UK gas customers – almost 40% lower compared to coal and 60-70% lower compared to electrolysis.

Similar cost curves can be developed for ammonia based on the key performance data given in Table 3.26.

	ATR	ELECTROLYSIS	GASIFICATION
Energy efficiency (%-HHV)	71	62.2	49.2
CO ₂ footprint	20.6	0	65.1
Specific cost (£/kW _{HHVH2})	1,301	1,902	2,453

Table 3.26: Summary of key performance data for large scale ammonia production

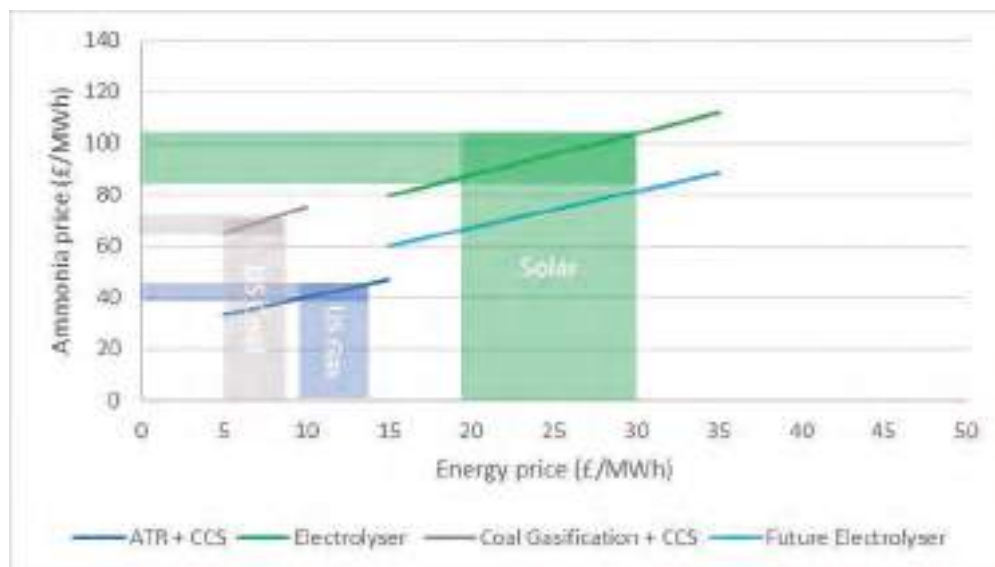


Figure 3.39: Cost curves for ammonia production technologies

The green bar is potential future solar electricity prices, blue is future US gas prices and grey is low cost coal prices.

Import of ammonia to UK can offer a longer-term option for hydrogen supply to the heating market. The analysis shows that a location like the US, where gas is cheap and access to potential CO₂ storage sites exist, the natural gas route offers the lowest cost alternative with a price of £45-50/MWh including transport to UK. In the longer-term green ammonia from solar power could be delivered at £70-120/MWh.

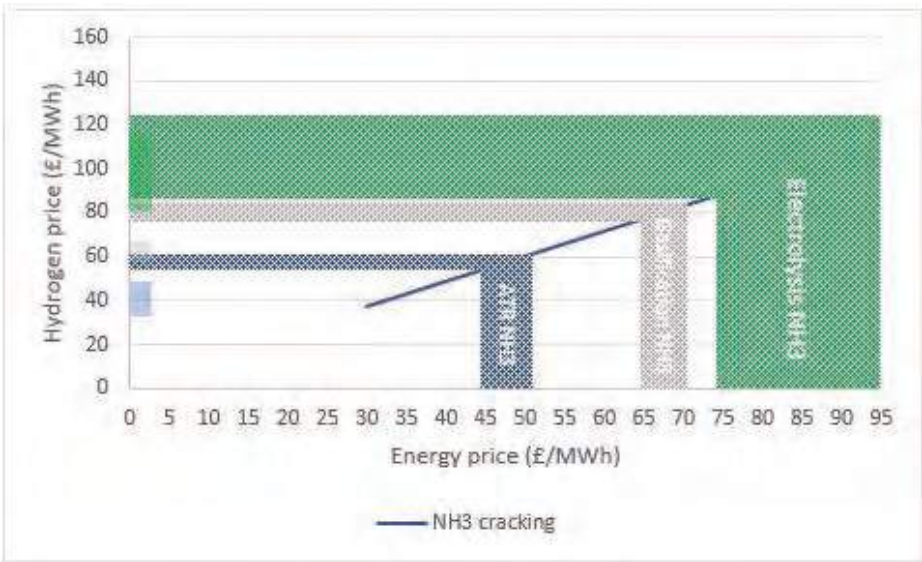


Figure 3.40: Hydrogen production cost from converting ammonia originating from three sources: natural gas (ATR+CCS), coal (gasification+CCS) and solar (electrolysis)



Assuming an additional £30/tonne for transport, i.e. £5/MWh ammonia converted to hydrogen in UK this will result in a hydrogen cost of above £80-120/MWh for green hydrogen from solar, which is higher compared to domestic hydrogen production from electrolysis at around £60-110/MWh (see Figure 3.40) hydrogen originating from natural gas based ammonia is about £55-60/MWh, which is about 40% higher compared to the direct hydrogen production alternative from ATR with CCS in UK.

3.8

Inter-seasonal storage technologies

For H21 NoE it is expected that inter-seasonal hydrogen storage will be required in the overall hydrogen supply concept.

In the H21 LCG report storage of compressed hydrogen in salt caverns was described in detail with an estimated working capacity of 855,000 MWh based on a 7 x 400,000 m³ cavern design operating at a 200 barg maximum pressure. Therefore, an extrapolation to H21 NoE requirements indicates a need for approximately 90 x 400,000 m³ caverns. This is equal to the number of caverns located at Etzel in Germany, which at 6 km² is, one the biggest sites of salt caverns in Europe (see Figure 3.41).



Figure 3.41: Etzel area – yellow circle highlights gas storage surface facilities

Since access to deep high-quality salt caverns in the UK is limited to the North-East other options for potential inter-seasonal storage have been examined, these include:

- Depleted hydrocarbon fields
- Ammonia
- Liquid hydrogen

In this section the different inter-seasonal storage method for H21 NoE will be presented with a specific focus on ammonia and salt caverns as the most mature. In the study an inter-seasonal hydrogen storage volume of approximately 1.5TWh has been used as a basis with a 30 days withdrawal rate equal to 2 GW. This is considered as typical size for one salt cavern hydrogen storage site and can be scaled up by developing multiple sites.

3.8.1 Introduction to underground storage

The main large-scale (TWh/GWh) storage methods used in the UK today are:

- Saline aquifers
- Depleted oil/gas fields
- Salt caverns

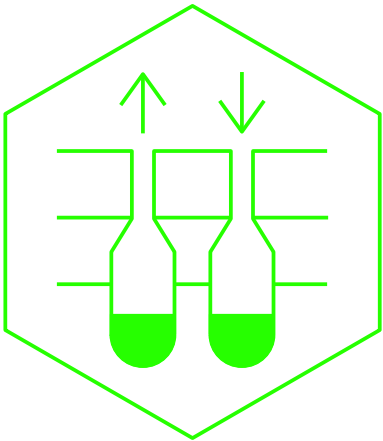
Saline aquifers and depleted oil/gas fields are mainly used for large scale storage, i.e. billion cubic metre (bcm) class. Due to limitation in cycle rate, i.e. the frequency and speed at which the gas is injected and extracted, these are mainly used for strategic/seasonal storage. Salt caverns are mainly used in the gas industry for multi cycle purpose with high injection and withdrawal rates²⁷. Further, salt caverns are recognised for being very gas tight and therefore well suited for hydrogen storage. There are examples of large-scale hydrogen storage in salt caverns. These are listed in Table 3.27.

PARAMETERS	CLEMENS DOME (USA)	MOSS BLUFF (USA)	SPINDLETOP (USA)	TEESSIDE (UK)
Operator	ConocoPhillips	Praxair	Air Liquide	Sabir
Year of operation	1983	2007	2017	1972
Volume (m³)	580,000	566,000	906,000	3 x 70,000
Depth (m)	1,000	1,200	1,340	365
Pressure range (bar)	70-140	55-150	70-200	45
Capacity (MWh)	81,000	123,000	274,000	27,000

Table 3.27: Parameters of hydrogen storage in salt caverns in USA and UK²⁸

All current hydrogen caverns are used by the chemical industry, which has high safety and quality standards for purity of hydrogen gas, demonstrating technical feasibility and integrity. The size and shape are comparable to state-of-the-art natural gas caverns, and techniques and methods for preparing a salt cavern for hydrogen are similar to those for natural gas.

The technical installation might not be comparable to the safety regulations in different regions. For example, in Europe there is a requirement for additional production tubings and subsurface safety valves²⁹.



27 Le Fevre, C., “Gas Storage in Great Britain”, Oxford Institute of Energy Studies, 2013
28 Letcher, T. M., “Storing Energy - with Special Reference to Renewable Energy Sources”, Elsevier, 2016
29 DBI GUT, “The effects of hydrogen injection in natural gas networks for the Dutch underground storages”, for Netherland Enterprise Agency, 2017

3.8.2 Hydrogen storage in salt caverns

This section describes a technical concept for large scale storage of hydrogen salt caverns for H21 NoE.

Working capacity

When developing the concept for hydrogen storage, although based on the same principles as natural gas, some differences needed to be considered. The following parameters need to be established in determining a suitable site:

- Hydrogen is a lighter molecule than natural gas with lower volume density (3 to 3.7 times lower) at the same pressure. In [Figure 3.42](#) the volumetric energy densities are shown as functions of pressure. This is the basis for determining the working capacity for the underground storage.
- Working capacity is defined by the stored volume between maximum and minimum pressure. Gas volume below minimum pressure is called cushion gas. Cushion gas is the amount of gas required to maintain the integrity of the cavern.

Example: A salt cavern hydrogen storage of 400,000 m³ with a maximum pressure of 200 bar (hydrogen at 200 bar has an energy density of 570 kWh/m³) and minimum pressure of 100 (hydrogen at 100 bar has an energy density of 300 kWh/m³) would have:

Working capacity: $400,000 \times (570 \text{ kWh/m}^3 - 300 \text{ kWh/m}^3) = 108,000 \text{ MWh}$

Cushion gas: $400,000 \times 300 \text{ kWh/m}^3 = 120,000 \text{ MWh}$
the amount of gas which must always be retained in the cavern (effectively part of the cavern establishment CAPEX costs)

- To maximise the storage capacity, higher maximum and lower minimum operating pressures are key operational requirements
- Standard guidelines for salt caverns is that the internal pressure must not exceed external pressure, i.e. pressure exerted on the cavern by the surrounding ground (lithostatic or overburden pressure). Exceeding this lithostatic pressure would result in tensile stresses and may fracture salt. Typically, a maximum operating pressure 10% below lithostatic pressure is reasonable for an adequate safety margin. Actual limitations for each cavern will depend upon Mechanical Integrity Tests (MIT) performed after completion. A typical overburden gradient of 0.2-0.22 bar/m is commonly assumed, and a typical maximum operating pressure is given by 0.17-0.2 bar/m. These were used in this study based on the extensive experience of cavern operators in the industry. For a salt cavern at 1,500 metre depths the maximum pressure is 255-300 bar.

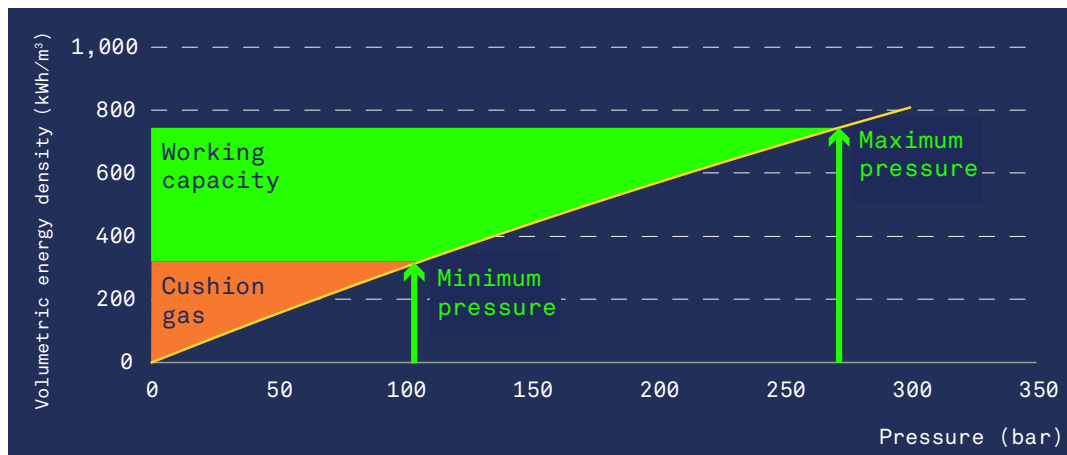


Figure 3.42: Hydrogen volumetric energy density at 20°C based on National Institute of Standard and Technology – US Department of Commerce

Salt cavern design

Designing and building cavities uses well established methodology. The following parameters need to be defined to enable the cavity design to be established:

- The cavity shape/geometry and adequate design pressures (minimum and maximum)
- The location of the cavity, i.e. the well pattern, the depth of the cavities, pillars (P/D ratios), distances to cap-rock and bedrock, distances to major geological faults etc.
- The planned maximum operating pressure in relation to the overburden pressure
- The minimum operating pressure

Increasing the density of caverns in a formation will increase the differential stress in the salt formation. This stress results from the difference between the in-situ overburden pressure on the salt formation and the operating pressure within the caverns.

Solution mining or leaching is used to create a void or cavity in an underground salt formation for storage purposes. This cavity is created by the dissolution of salt in fresh water or unsaturated brine injected into the formation under controlled conditions. The process is illustrated in Figure 3.43.

The injection of fresh water/unsaturated brine and removal of saturated brine is achieved via two pipe strings. These are suspended concentrically below the production casing string. Water is injected into the developing cavity through the wash string. Brine is removed through the solution mining string. The depths and position of these strings relative to each other and to the cemented production casing shoe are based on the desired shape and geometry of the storage cavern and on the solution method employed.

An essential part of the solution mining operation is the placement and maintenance of a blanketing material (nitrogen, natural gas, etc.). This fills in the annular space between the production casing and the suspended wash and solution-mining strings.

The blanketing material prevents the removal of the salt-cement seal around the permanently cemented casing strings, limits the upward growth of the cavern and helps to control the shape of the cavern. The blanketing material must be insoluble in salt, immiscible with water and brine, and its density less than that of water.

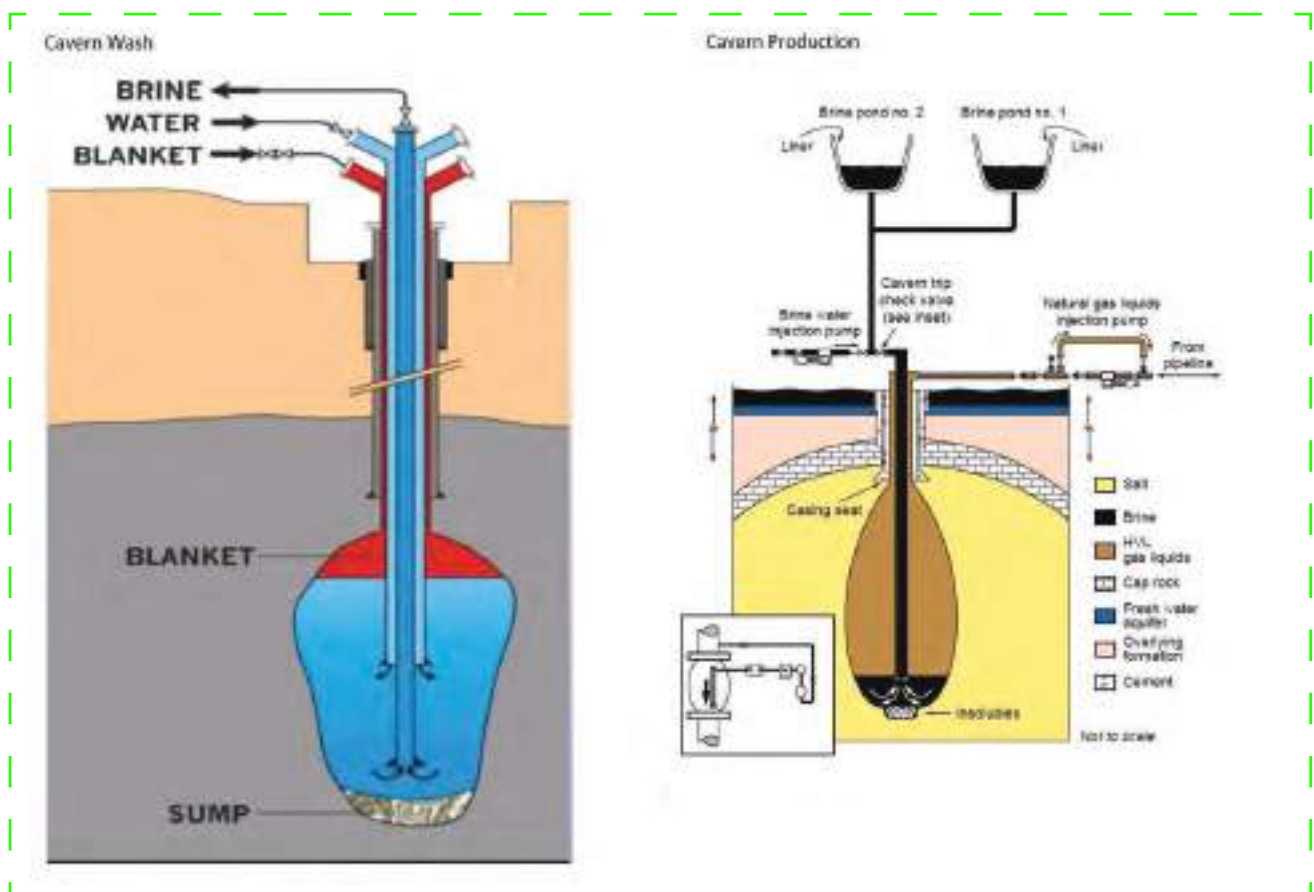


Figure 3.43: Water is injected and dissolves the salt into a brine – the brine is pumped out and the void is filled with a blanket (nitrogen) to avoid further growth in the leached area

The development of the cavity during leaching is controlled by the following factors:

- Water injection rate, which determines the rate of cavern growth
- Blanketing material and water interface level, which protects the cemented production casing shoe, establishes the upper limit of cavern growth, and can exert an influence on horizontal cavern growth.
- Water injection and brine removal points. Locations in the well relative to each other and the blanket-water interface, determine the cavern shape along with the method of circulation involved

Based on experience from the gas storage project in Aldbrough, the storage parameters summarised in Table 3.28 have been selected for a standard hydrogen salt cavern storage site with one surface processing unit:

KEY DESIGN PARAMETERS	UNIT
Cavern size	300,000m³
Depth	1,700-1,800metre
Number of caverns per surface facility	10
Minimum operating pressure	85bar
Maximum operating pressure	275bar
Maximum daily differential pressure	10bar
Gas volumes	
Cushion gas (10 x 300,000 x 260kWh/m³)	780,000MWh
Working gas	1,440,000MWh
Total gas (10 x 300,000 x 740kWh/m³)	2,220,000MWh
Withdrawal capacity in 30 days	2,000MW c.13.5Msm³/d
Injection capacity in 60 days	1,000MW c.6.75Msm³/d

Table 3.28: Key design parameters for a hydrogen salt cavern

Surface processing facility

The surface facility comprises the necessary equipment to inject and withdraw hydrogen to and from the salt cavern. Compression to required cavern pressure would be done in a reciprocating compressor. However, due to the temperature increase during compression, cooling of the hydrogen will be needed in order not to exceed maximum operating temperature of the salt cavern and wells and enable higher volumes of hydrogen storage. The compression and injection operation will be carried out during spring and summer when the hydrogen storage would need to be filled for the following winter.

During winter, hydrogen will be withdrawn from salt caverns via control valves regulating the required flow and meeting HTS pressure. Since the pressure in the salt cavern is significantly higher compared to HTS, hydrogen will be expanded via the control valve. During the expansion of hydrogen temperature will increase and cooling is needed so the HTS design temperature is not exceeded.

In addition, metering of injected and withdrawn hydrogen will be carried out.

In [Figure 3.40](#) the principles of a hydrogen storage surface processing unit are presented.

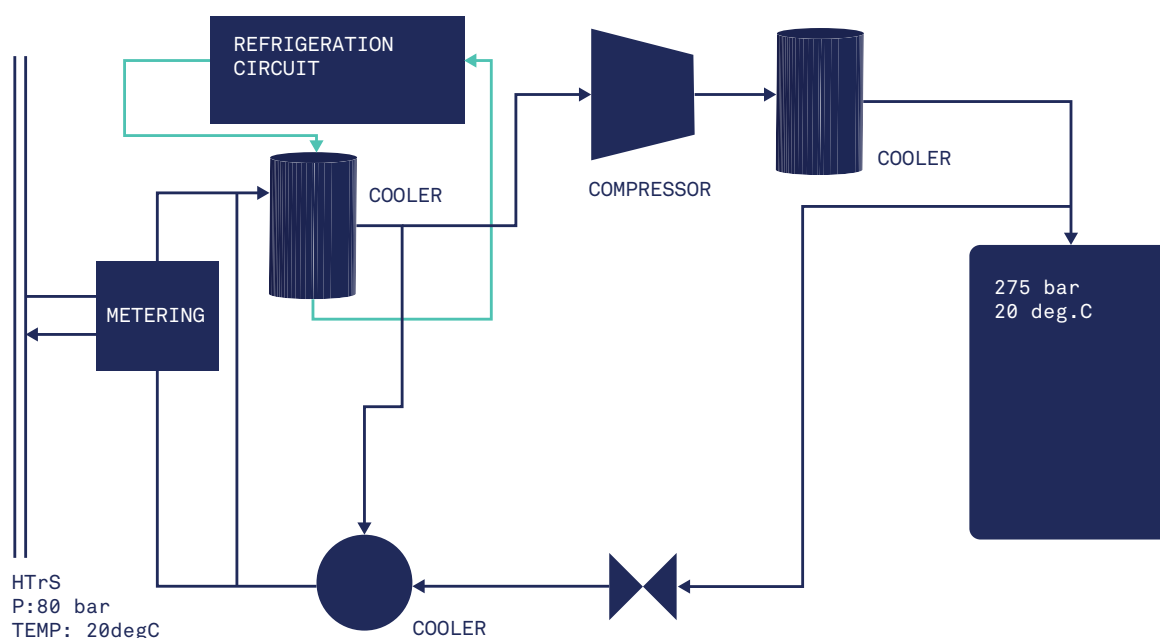


Figure 3.44: Hydrogen storage surface processing unit

Based on the key design parameters given in [Table 3.26](#) the typical capacities of the main equipment have been estimated and are given in [Table 3.29](#).

CAVERN SURFACE FACILITY	
Compression	3 x 8,000kW reciprocating compressor with motor drive
Ammonia refrigeration	3 x 750kW ammonia compressor and refrigeration circuit units
Air coolers	3 x 150kW air coolers

Table 3.29: Hydrogen storage surface facility equipment

Capital cost

CAPEX estimates have been developed based on experience from the existing UK operational gas storage project in Aldbrough in relation to subsurface work. In addition, quotation from suppliers on critical equipment, compressors, heat exchangers, wells, metering and valves have been used to establish an estimate for the surface unit. To summarise, a typical hydrogen storage project using salt caverns will comprise the following elements:

- Site service: Site preparation works, site facilitating works, energy supplies
- Subsurface work: Drilling and workover operations, snubbing operations, leaching plant and leaching operation
- Water supply and brine disposal
- Surface processing unit
- Project management and engineering services
- Misc.: Environmental, insurance, decommissioning

There will be a need for approximately 780 GWh of cushion gas. Assuming the cost is equal to the cost of the natural gas required to produce the hydrogen cushion gas the cost can be estimated as follows (assuming a natural gas price of £23/MWh (BEIS 2035 forecast)):

780,000 MWh/74.7% (efficiency of Hydrogen Production Facility, [see Section 4.0](#)) x £23/MWh = £24,000,000. The total CAPEX is estimated to £301,000,000 with following break-down excluding cost of land and permitting etc.

WBS	(£M)
Site service	22
Subsurface work	80
Coastal	10
Surface processing unit	126
Management	20
Miscellaneous	14
Contingency (10%)	29
Sub total	301
Cushion gas	24
TOTAL	325

Table 3.30: CAPEX breakdown of hydrogen salt cavern

A total cost £325m equals a specific storage cost of £225.7/MWh working capacity. In the H21 LCG the storage was estimated to £289m for 855,000 MWh working capacity equal to £338/MWh.

Operating and maintenance cost is assumed to be 4% of total CAPEX, i.e. 0.04 x 301,000,000 = £12,000,000. This will be further detailed in [Section 4.0](#) for the H21 NoE full design.

The area required for the surface facility including safety zones is about 125 x 150 metres.

3.8.3 Hydrogen storage in depleted oil/gas fields and aquifers

Storing hydrogen in large volumes in depleted oil/gas fields or aquifers for strategic and seasonal purpose presents an interesting option in combination with salt caverns.

Hydrogen storage has been performed for town gas (containing 50% hydrogen by volume) storage in countries like Germany, France, Belgium, Czechoslovakia and Poland until 1990 to guarantee the supply. At that time, 8 of the 13 aquifer reservoir stores, 3 of the 15 depleted gas field stores and 2 of the 19 salt cavern stores in Germany were operating with town gas. Currently, all of them are either operating with natural gas or are closed.

Natural gas reservoirs normally do not contain hydrogen, but hydrogenous town gas was indeed stored in porous formations. No problems regarding the sealing/tightness and hydraulic integrity of the cap rock was reported at that time. Nonetheless, hydrogen storage under the planned operational conditions is a novelty due to its specific properties like molecular size and weight. The geological tightness of a porous formation cap rock against hydrogenous gas depends on the cap rocks ability to withstand gas infiltration both mechanically and hydraulically.

Permeability needs to be considered as well as diffusion and solution processes. A decisive factor is also the water saturation of the cap rock, differing from reservoir to reservoir. Hydrogen must dissolve into the formation water of the cap rock when diffusing through the cap rock. The diffusion is dependent on the concentration gradient, where the maximum concentration of a gas component is dependent on its solubility in water. The solubility of hydrogen in water is lower than in methane, but the diffusion ability of hydrogen is four times higher. Large-scale methane losses by diffusion are not expected in former natural gas reservoirs (which are now used for storage), because the formation water in the cap rock has already been saturated with methane over geological periods. Due to that, there is only a very small gradient. For storage of hydrogen, this gradient exists in the beginning and its value depends on the hydrogen content in natural gas.

Even though diffusion is a slow process, small initial losses need to be considered at the beginning of hydrogen storage. Those losses appear until the formation water in the adjacent cap rock is saturated with hydrogen. Hydrogen losses in the range of 2% over the life cycle of a Underground Store (UGS) have been reported. A physically different process is the permeability of the cap rock. Here it must be considered that the pore volume of the cap rock is saturated with water and the sealing is caused by capillary pressure, which should not be exceeded. The influence of hydrogen on capillary pressure has not been reported yet as it is assumed it depends on the interfacial tension between water and hydrogen, which is like methane and water systems.

During storage of hydrogen, there is the risk to trigger geo-chemical reactions with rock minerals and reservoir fluids. Those reactions could on the one hand lead to energy losses and, on the other hand, to further undesirable effects. Reactions with rock minerals could lead to damage in the rock and mineral structure, resulting in the alteration of crucial reservoir properties like pore volume and permeability.

The biological integrity will also have to be considered. Some microbes can metabolise hydrogen and therefore reduce the hydrogen content and quality of the gas and reduce reservoir permeability. The microbial status of a potential storage site and how this adapts to a hydrogen environment will need to be investigated.

In the case of natural gas underground storage, the secure injection and production cycle always must be guaranteed. Therefore, the main consideration is that the occurrence of uncontrolled gas leakages in all parts of the borehole completion and surface facilities must be avoided during all parts of the operation phases. This process, also known as borehole integrity, is the total of all procedures that are ventured during planning, drilling and usage of an underground storage well, to guarantee the injection and production of fluids as well as the leak tightness. Therefore, a lot of technical, operational and organisational procedures and solutions are needed.

It is recommended that the former gas storage at Rough is evaluated for potential hydrogen storage.

3.8.4 Ammonia as inter-seasonal hydrogen storage

Ammonia storage is a mature and well-established technology used at large scale worldwide. Significant experience has been gained in more than 60 years to develop best practices and safety standards. Since ammonia is a toxic substance, large scale ammonia storage is only expected to be done at industrial sites. This is where skilled and trained personnel are present and safe operation and maintenance can be achieved. Ammonia leakages from industrial ammonia storages are low. In the UK ammonia emission from industrial production only accounts for 0.3% of total ammonia emissions, agriculture being far the largest ammonia emitter³⁰.

Ammonia is normally stored in liquid form under atmospheric conditions and a temperature of -33°C . The high energy density of 6,200 kWh/tonnes or 4,300 kWh/m³ allows storing of large volumes of energy. Since ammonia is extracted from the storage tank by a pump, the efficient storage volume is almost 100%.

A typical ammonia storage tank is single or double walled and refrigerated. The largest built are 50,000 tonnes net capacity of approximately 50 metres in diameter and 40.5 metre height (Figure 3.45). Storing the same amount of energy as the ten-cavern hydrogen storage sites would require five tanks of 50,000 tonnes.



Figure 3.45: Ammonia tank of 50,000 tonnes for Qatar Fertiliser Company (QAFCO) ammonia and urea expansion project in Mesaieed, Qatar. Courtesy: CB&I

Single wall tank: This comprises of a flat bottom tank with a dome roof designed to withstand the stated conditions. Tanks of this type will have external surfaces insulated to minimise heat gain from the surroundings.

In addition to the ammonia storage, an ammonia synthesis process facility and an ammonia cracker facility would be required to meet the overall objective of delivering hydrogen (see Figure 3.46).

Double wall integrity tank: This tank consists of double wall construction with both tanks designed so that in the event of failure of the inner shell the ammonia will be contained by the outer shell. The inter-space between the tanks will contain only cold ammonia vapour.



Figure 3.46: Ammonia-to-hydrogen inter-seasonal storage concept

³⁰ DEFRA, "Ammonia in UK", 2,002

3.8.5 Ammonia storage concept design



Based on the similar design basis to the hydrogen salt cavern presented in [Section 3.7.2](#) (storing 1,490,000 MWh for 30 days), the ammonia storage would have the key parameters given in [Table 3.31](#).

KEY DESIGN PARAMETERS	UNIT
Ammonia cracker unit based on 30 days extraction	2,067 MW
Ammonia tanks	5
Ammonia tank size	55,200 tonnes
Ammonia synthesis unit based on 60 days production	1,198 MW or 4,600 t/d
Hydrogen consumption in ammonia synthesis	1,344 MW
Nitrogen consumption in ammonia synthesis	96 t/hr

Table 3.31: Key design parameters of the ammonia-to-hydrogen inter-seasonal storage concept

Ammonia cracker unit

Based on case 2 in [Table 3.22](#) in [Section 3.6.5](#), approximately 2,395 MW of ammonia is needed to produce 2,070 MW of hydrogen equal to 1,725,000 MWh or 276,000 tonnes ammonia in 30 days. In addition, 25.1 MW of power import is required.

Ammonia storage

Ammonia storage is based on a double-wall integrity design including refrigeration systems for ammonia vapour liquefaction. Storage tank size of 55,000 tonnes has been assumed. A drain system is included to drain vessels for maintenance. The liquid from drain lines are sent to the vent drum where evaporated vapour ammonia from the top is released to the flare header. The flare stack is normally 30-40 metre high.

Ammonia produced for storage is based on 60 days production capacity, i.e. 1,198 MW (191.7 tonnes/hr or 4,600 t/d). To achieve this production capacity, about 1,344 MW of hydrogen together with 96 t/hr of nitrogen is needed. Prior to the ammonia synthesis the hydrogen will enter a clean-up unit where the remaining CO and CO₂ will be converted into methane using a catalytic reactor.

Since the ammonia synthesis is exothermic, 230 t/h of medium pressure steam at 55 bar and 275°C will be generated to run steam turbines for compression. The total power produced in the turbines will about 42 MW and will be distributed to the nitrogen compressor (10 MW), hydrogen compressor (2 MW), recirculation compressor (5 MW) and refrigeration compressor (20 MW) resulting in a net surplus of 5 MW.

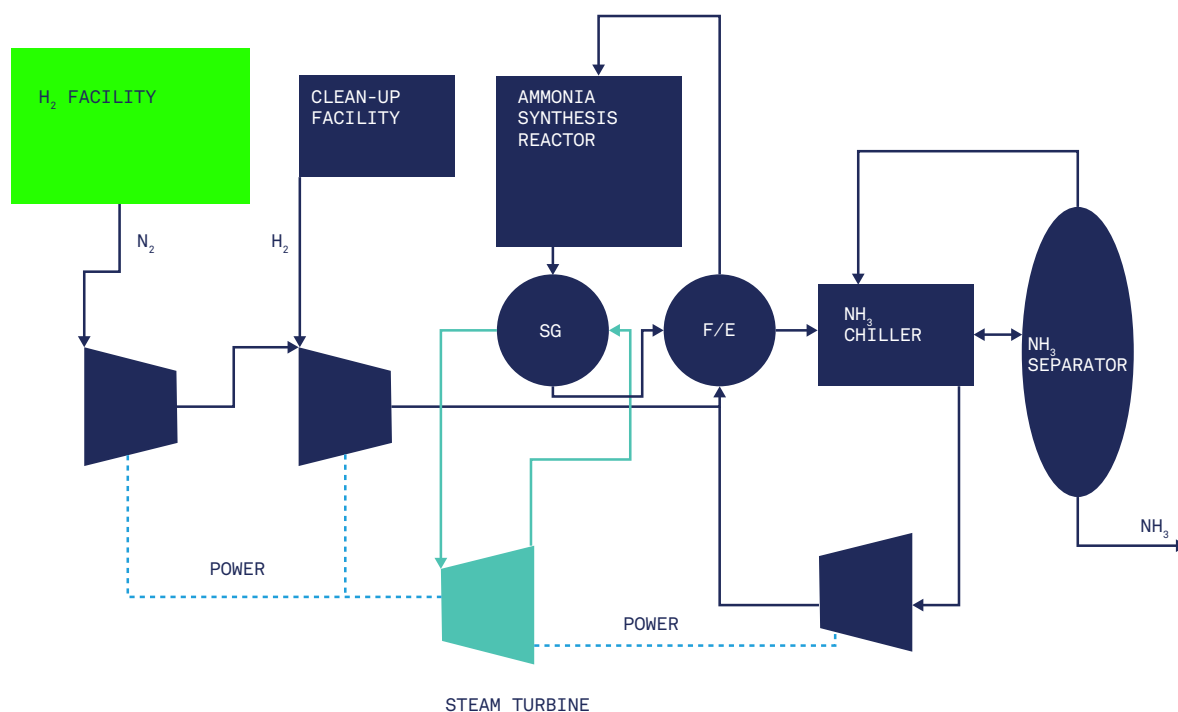


Figure 3.47: Ammonia synthesis unit

(1,935,600-1,490,000)/74.7% (efficiency of Hydrogen Production Facility, [see Section 4.0](#)) x £23/MWh = £13,720,000

Operating and maintenance cost is assumed to be 4% of total CAPEX, i.e. $0.04 \times 1,247,000,000 = \text{£}49,880,000$.

The ammonia to hydrogen inter-seasonal storage facility will most likely occupy the same plot area as a completed ammonia production facility as described already in this section.

(£M)	AMMONIA SYNTHESIS UNIT	AMMONIA STORAGE UNIT	AMMONIA CRACKER UNIT
Equipment	121		123
Bulk	99		101
Indirects	19		20
Construction	92		94
Home office	6		6
CMT	13		13
Other	4		4
EPC (£M)	354	160	362
Owners cost	30	10	50
STL project managent	18	10	18
Insurances	11	5	11
Contingency	83	37	88
TOTAL PROJECT COST	496	222	529

Table 3.32: CAPEX breakdown of ammonia to hydrogen inter-seasonal storage concept

However, some options exist to improve the concept:

- Producing ammonia in the summer season from surplus hydrogen from the Hydrogen Production Facility (say April to October). This would allow a much slower 'fill' of the ammonia storage with an associated reduction in the ammonia production facilities capacity by circa 72%. This would then have an additional CAPEX benefit with a saving of approximately £190m.
- Import of ammonia for strategic storage/ inter-seasonal storage in combination with hydrogen storage in salt caverns
- Utilise curtailed electricity as heating source for ammonia cracking

However, it is clear from this analysis that the advantage of an ammonia to hydrogen storage facility is primarily a concept for strategic low cycle inter-seasonal storage and it becomes an expensive concept when requiring the same type of fast output/input rate as the salt cavern.

3.8.6 Liquid hydrogen as inter-seasonal hydrogen storage

Using liquid hydrogen as inter-seasonal storage is illustrated in Figure 3.48 and will be based on the following main principles:

- Pre-treatment of fuel grade hydrogen from ATR using PSA technology
- Liquefaction of hydrogen
- Storage of hydrogen
- Pumping and regasification of hydrogen

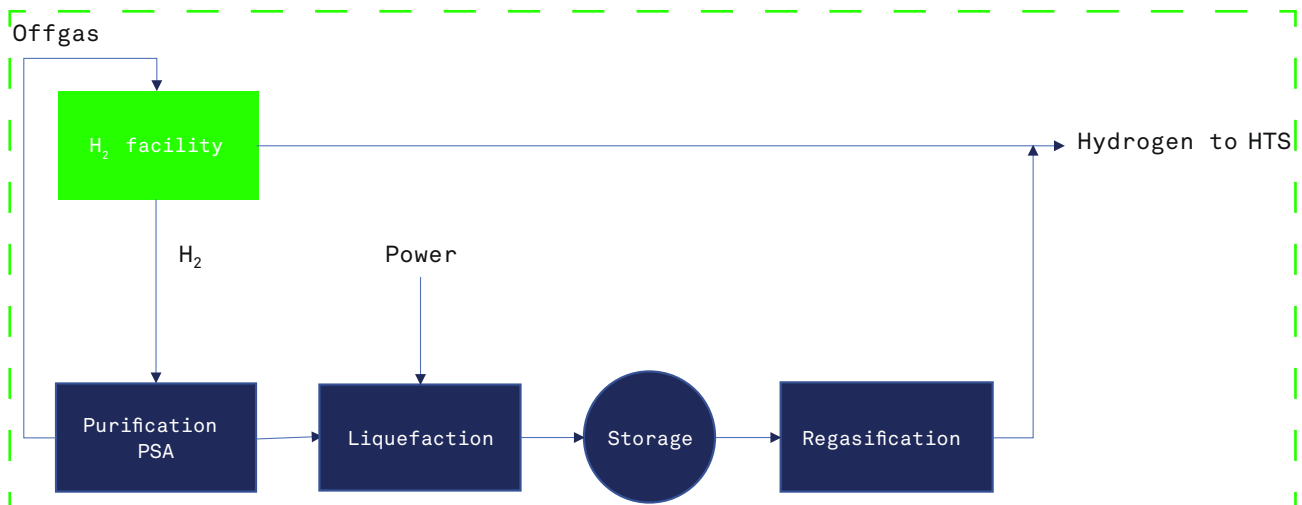


Figure 3.48: Inter-seasonal storage of liquid hydrogen

Liquid hydrogen is an established and proven technology but only in operation at small scale:

- Liquefaction of 5-15 tonnes per day equal to 8.2-24.8 MW (HHV)
- Liquid hydrogen storage of 3,000 m³ and with a density of 70 kg/m³. This equals 8.27 GWh (HHV)

Liquefaction of hydrogen requires significant energy in the form of power, typically above 10 kW/kg H₂, or approximately 25% of the energy content in the hydrogen. It is however expected that technology improvement and scale-up could result in lower power consumption in the range of 7.5-9.0 kW/kg³¹. Current referenced scale of liquefaction technology is 15 t/d; however, 200 t/d is considered achievable leading to some cost reduction.

Purification of the fuel grade hydrogen from the ATR facility would be based on Pressure Swing Adsorption, the conventional method in the hydrogen industry. Large scale PSA normally has a hydrogen recovery rate of 90%. Therefore, 10% of the fuel grade hydrogen leaves as off gas at low pressure and must be returned to the ATR facility for use in the fired heater and boilers. It can be assumed that this will replace the current hydrogen fuel and therefore not impact the energy balance of the hydrogen plant.

Based on the inter-seasonal storage assumptions (1.5 TWh of storage, 1 GW injection capacity and 2 GW withdrawal capacity), it is possible to establish the main design parameters for a LH₂ inter-seasonal storage.

³¹ Klaus Ohlig, Linde Group, "Latest developments and outlook for hydrogen liquefaction technology", 20th World Hydrogen Energy Conference 2,014

Total cost is estimated to be £6,289 m or £4,193/ MWh stored. As seen from the main cost driver is the cost of storage tanks and liquefaction facility.

£M	PURIFICATION (PSA)	LIQUEFACTION	STORAGE	REGASIFICATION
Design	313,000 Nm ³ /h	3 x 200 t/d	180 x 3,000 m ³	1,220 t/d
Equipment	30	350		40
Bulk	25	287		33
Indirects	5	57		7
Construction	23	267		31
Home office	1	17		2
CMY	3	37		4
Other	1	11		1
EPC (£M)	88	1,027		117
Owners cost	5	142		10
STL project management	4	51		6
Insurances	3	31		4
Contingency	21	250		29
TOTAL PROJECT COST	122	1,502	4,500	165

Table 3.33: Capital cost breakdown for a liquid hydrogen inter-seasonal storage concept

Withdrawal capacity of 2 GW equals a regasification capacity of 1,220 t/d. Power consumption would be less than 1% of the hydrogen energy content, i.e. 20 MW. Regasification of hydrogen is assumed to be achieved by heat supplied from water or air.

Storage of 1.5 TWh net capacity equals 38,000 tonnes of hydrogen requiring 542,500 m³ of liquid hydrogen storage capacity. This would be done in 180 spherical 3,000 m³ storage tanks. It should be noted that 0.2% of the hydrogen will boil-off daily, or 50% of a full tank in one year. From operational experience, this means that the storage should be filled as close as possible to the winter season. If a mild winter occurs, where there is little need for additional hydrogen, the hydrogen plant must be turned down and hydrogen extracted from the storage. However, since the inter-seasonal storage must accommodate capacity for a peak year, strategic reserves will be needed. Due to the boil-off rate, it is unlikely LH₂ can serve that purpose.

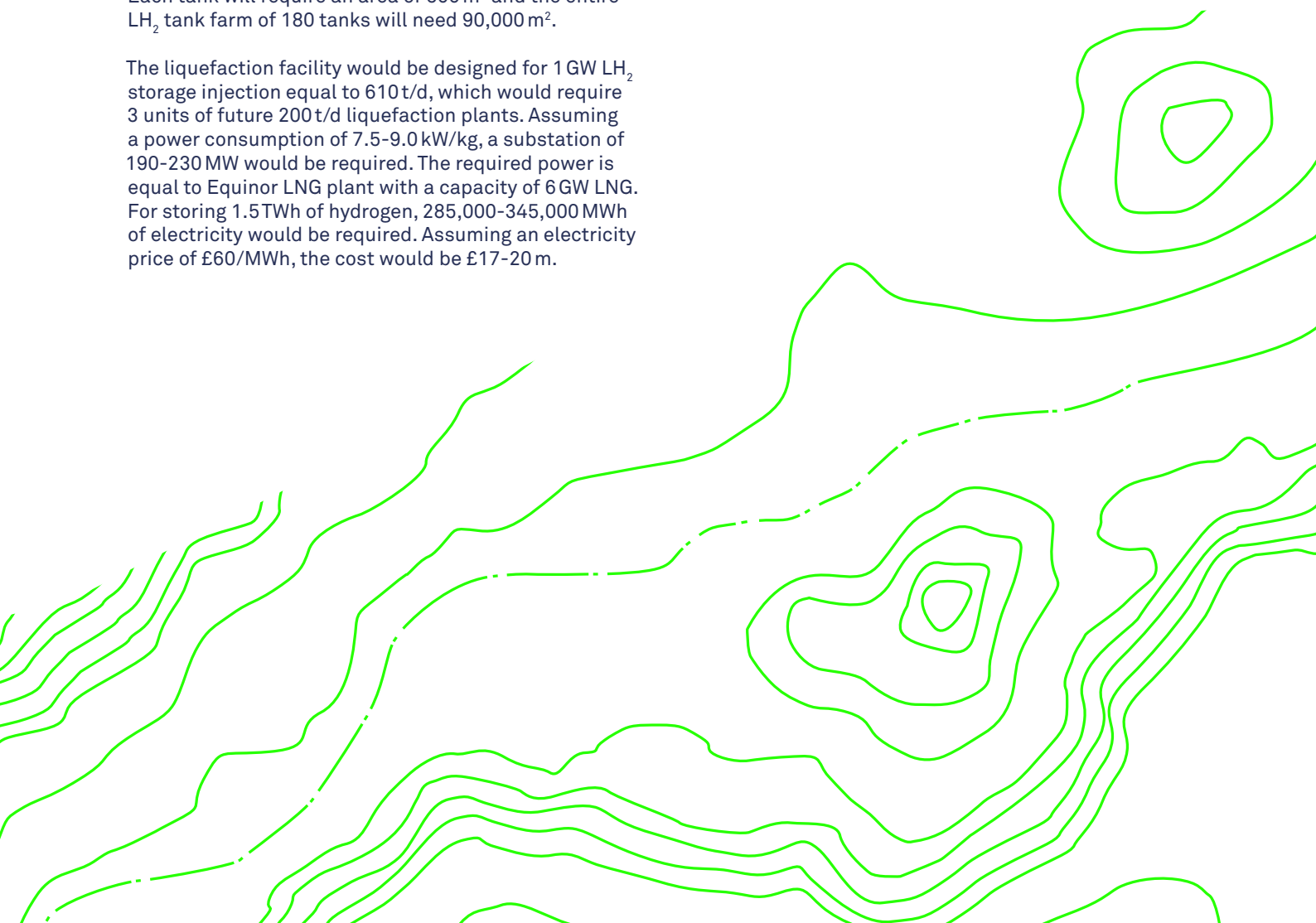
The dimension of the tank is 18 metres in diameter. Each tank will require an area of 500 m² and the entire LH₂ tank farm of 180 tanks will need 90,000 m².

The liquefaction facility would be designed for 1 GW LH₂ storage injection equal to 610 t/d, which would require 3 units of future 200 t/d liquefaction plants. Assuming a power consumption of 7.5-9.0 kW/kg, a substation of 190-230 MW would be required. The required power is equal to Equinor LNG plant with a capacity of 6 GW LNG. For storing 1.5 TWh of hydrogen, 285,000-345,000 MWh of electricity would be required. Assuming an electricity price of £60/MWh, the cost would be £17-20 m.

Finally, the PSA purification would treat 313,000 Nm³/h of hydrogen from the ATR and producing 282,000 Nm³/h equivalent to 1 GW of hydrogen – 31,000 Nm³/h off-gas will be returned to the hydrogen facility and replace fuel for heaters and boilers.

A capital cost breakdown has been established based on the following:

- Budget price for PSA
- Extrapolated from LNG plant cost using power input as the main parameter
- LH₂ storage cost based on Equinor experience from the Viking Cruise project and cost and cost estimates from Equinor on-shore tank farm projects
- Re-gasification cost extrapolated from LNG regasification experience
- Installation factors, etc. based on Equinor's experience from onshore projects



3.8.7 Summary of inter-seasonal storage

Based on an inter-seasonal storage volume of approximately 1.5TWh as a basis with a 30 days withdrawal rate equal to 2 GW. A comparison of saline aquifers, ammonia, liquid hydrogen and salt caverns are presented in the following table.

	TECHNOLOGY READINESS LEVEL	EFFICIENCY	TOTAL CAPEX £/MWH
SALINE AQUIFER/ DEPLETED HYDROGEN CARBON	Unproven, needs investigation for example into re-use of the Rough storage field in the North Sea	Unproven	N/A
AMMONIA	Proven at the correct scale to be appropriate for inter-seasonal hydrogen storage considerations. However most appropriately for few cycles per annum at steady rate of injection and extraction.	Significant losses (circa 25% in total) in the production and extraction process.	830 (can be further optimised, see comments in Section 3.7)
LH ₂	Proven storage but only at small scale. Not currently appropriate for inter-seasonal hydrogen storage.	Significant losses in the refrigeration process and from 'boil off' losses associated with long term storage.	4,193
SALT CAVERN	Proven storage at scale with a long history of reliable, safe performance on a range of gases including natural gas, towns gas and hydrogen. Especially suited for both annual multiple cycles.	Very low efficiency losses mainly associated with compression requirement.	225.7

Table 3.34: Inter-seasonal storage summary

Inter-seasonal storage of hydrogen is critical for conversion of H21 NoE to 100%. Salt cavern storage is the most mature technology and offers the lowest cost for storing large volumes in TWh hour scale.

3.9

H21 NoE system considerations



In [Section 3.3 to 3.8](#) the potential large-scale hydrogen production and inter-seasonal hydrogen storage technologies have been presented and analysed.

This section will describe the entire system required to provide hydrogen for the H21 NoE area of conversion.

The following systems have been analysed:

- Offshore wind electrolysis-based hydrogen production in the UK
- Natural gas UK based hydrogen production using ATR and CCS
- UK located coal gasification plant with CCS
- Ammonia production in the US transported to the UK by ship and cracking into hydrogen in UK based on natural gas, coal and solar power

3.9.1 Assessment of water electrolysis for H21 NoE

The key driver to convert gas networks to 100% hydrogen is the overall carbon reductions of the system, and, in this context, the system design needs to be considered. The electrolyser process is not emitting any CO₂ directly but may lead to indirect emissions as a result of the electricity production. In Table 3.35 the carbon footprint of hydrogen produced from electrolysis with 70.9% efficiency is shown from different electricity sources.

ELECTRICITY SOURCE	CO ₂ FOOTPRINT ELECTRICITY (G/KWH)	CO ₂ FOOTPRINT HYDROGEN (G/KWH)
Coal power	700	987.3
Gas power	350	493.7
UK grid target of 100g/kWh	100	141.0
UK grid target of 50g/kWh	50	70.5
Wind or solar	0	0.0

Table 3.35: CO₂ footprint of hydrogen from electrolysis using different power sources

With a clear goal of achieving significant carbon reduction in heat only an electrolyser directly connected to a renewable power source seems attractive when considering these numbers. In the UK this would strongly suggest integration of electrolysers with access to huge offshore wind resources as well as a mature and growing wind industry.

At this stage it was considered adequate to use some global numbers for the net capacity factor and cost. It has also been assumed that the electrolysis load can follow the power output and there is no resultant efficiency or production penalty. It must be stated that this is most unlikely. An illustration of this value chain is shown in Figure 3.49.

The project team therefore assessed the potential for the UK to produce its green hydrogen from offshore wind electricity. For a project such as converting the NoE to hydrogen, it is assumed that a centralised electrolyser-based facility would be directly connected to the High Voltage Direct Current (HVDC) coming from the offshore wind farms. In the evaluation, the wind farm and power transmission components have not been fully assessed.

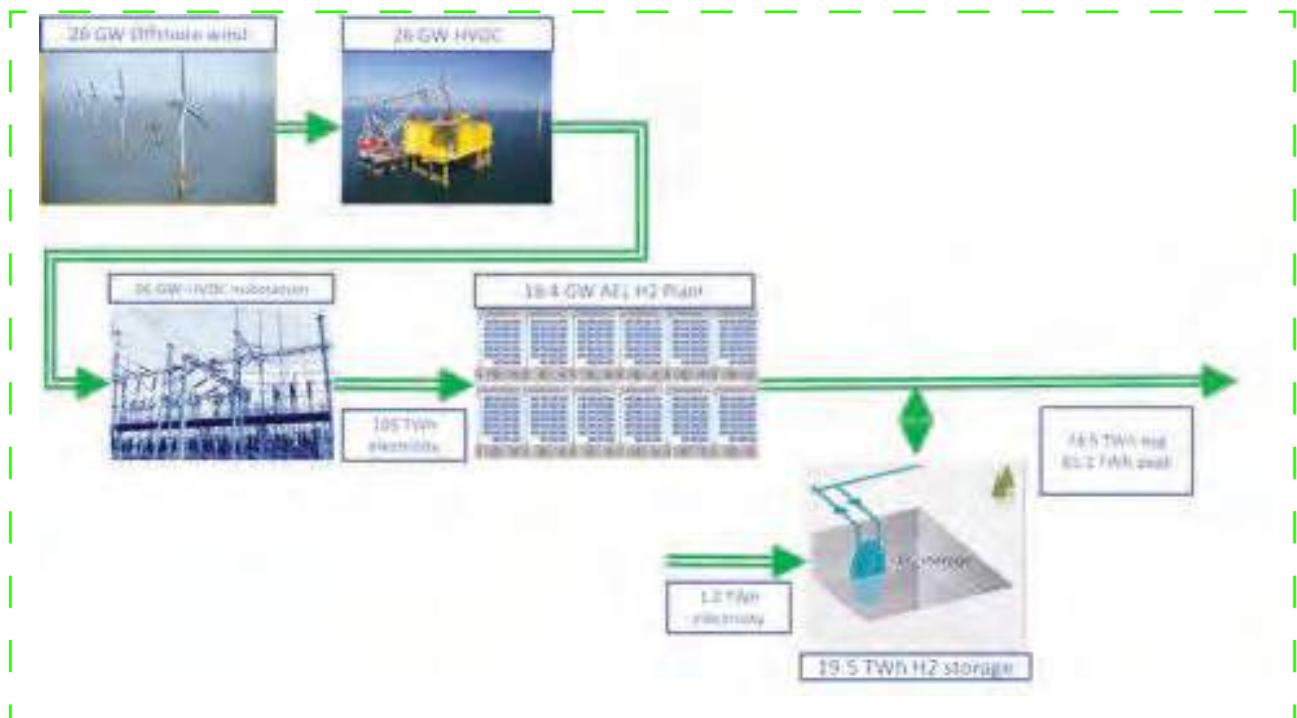


Figure 3.49: Offshore wind and water electrolysis value chain for the NoE

The NoE heat demand for the years 2013-2016 is shown in Figure 3.50 together with a wind energy production profile for the same period. The wind power supply was taken from an Equinor offshore windfarm and scaled to match the NoE demand over the period 2013-16.

The poor match between supply and demand is obvious, creating a need for a very large inter-seasonal hydrogen storage solution to secure the supply.

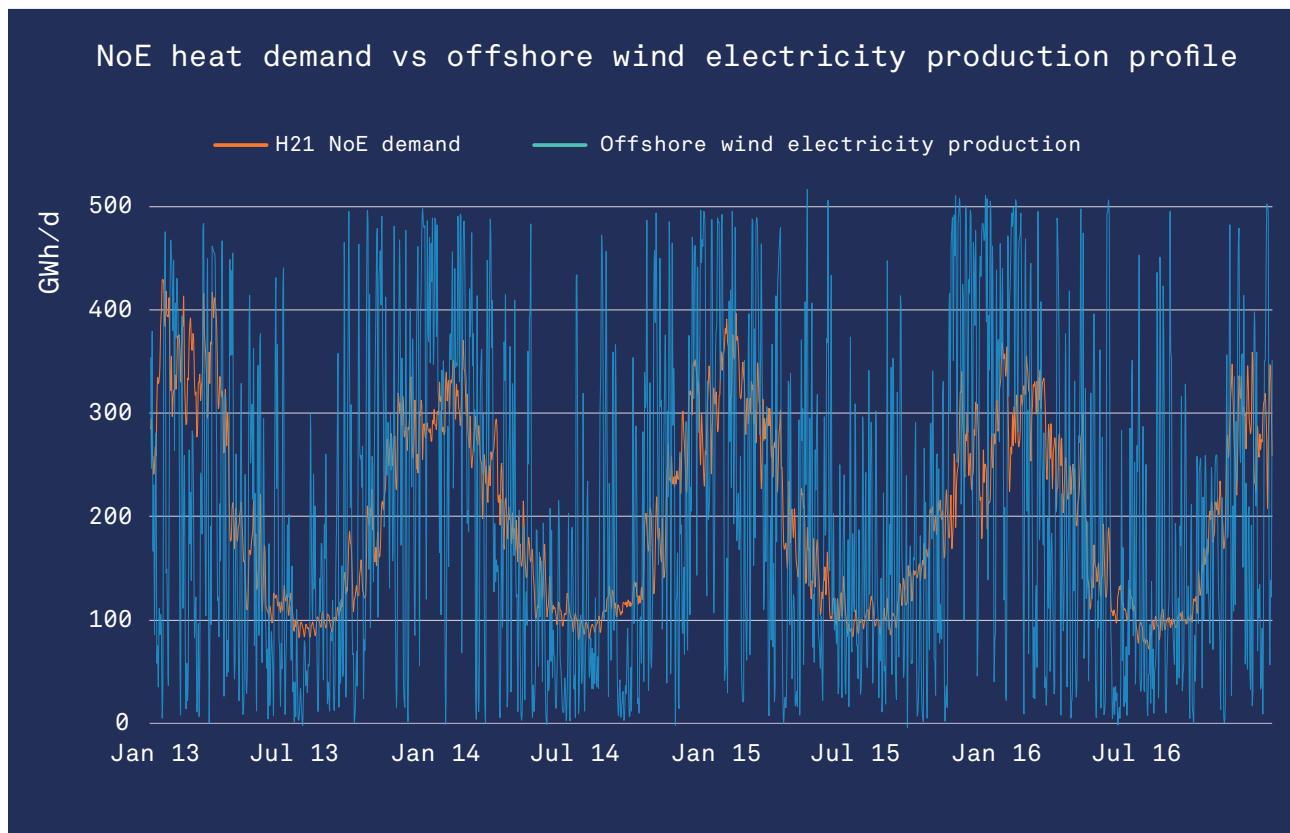


Figure 3.50: NoE heat demand versus offshore wind electricity supply profile

The supply chain was further analysed to scale the demand for the year 2013 to match the peak year demand of 85.12TWh quantified in [Section 2.0](#). The wind, water electrolysis and storage capacity were scaled to match this demand assuming an electrolyser efficiency of 70.9% (see [Section 3.3](#)). No transmission or conversion losses have been included. The results are given in [Figure 3.51](#), showing a required storage capacity of 19.5TWh of hydrogen equal to 130 caverns of 300,000m³.

26GW of installed offshore wind capacity and 18.4 GW of hydrogen production capacity is required to match this demand.

The demand for offshore wind corresponds to 5 of the current planned 4.8GW Doggerbank wind projects. If Doggerbank is built it would be the largest offshore wind project in the world covering a sea area of 2,267 km². Based on Doggerbank, the required area of the H21 NoE wind supply is more than 11,000 km². To put this into further context, we may look at the collected timeline of UK offshore wind projects.

An overview was developed by Renewable UK. This was in co-operation with the Offshore Wind Programme Board and The Crown Estate with the support of the Offshore Wind Industry Council (published in 2016). It reports a total wind power capacity of 5.15 GW in 2016. This was expected to double by 2020 and then double again by 2030, suggesting a total UK offshore wind output of at least 20.6 GW by 2030³². Hence, the total supply of offshore wind in the UK might be able to match the total NoE demand in 2035. But it is not realistic that all this supply could be directed to NoE.

The water electrolyser supply chain is immature when targeting a multi-GW regional deployment of hydrogen production technology from 2026 onwards. It is considered unlikely that the electrolyser market will be able to deliver at a scale for a project like H21 NoE conversion within the required time frame. The 18.4 GW electrolyser facility would need to have more than 10,693 Nel A-485 (the biggest referenced alkaline electrolyser) installed over the period 2026-2035 at more than 1,100 units per year. This is beyond the scale that traditional water electrolyser manufacturers can provide today.

Even if a much larger company like Thyssenkrupp succeeds with the deployment of their electrolyser concept, they would only be able to deliver 600 MW of capacity per year. It would take 40 years to cover the NoE demand with full allocation of supply capacity to this project.

If, however, the UK meets the target of 50 g/kWh in 2050 for the electricity production, 30% of the electricity could be purchased from the grid and the remaining produced directly from a renewable source.

This would achieve an overall CO₂ emission of 15 g/kWh which is comparable with the natural gas ATR hydrogen supply chain. This would reduce the size of the offshore wind park, electrolyser and storage by 30%. This could potentially reduce the cost of hydrogen by 10-15% and increase the robustness of the value chain as grid electricity import could be used to level out the peaks in the offshore power production.

Seasonal demand (orange), hydrogen production rate (blue) and seasonal storage reserves (green). Installed wind power capacity 26 GW

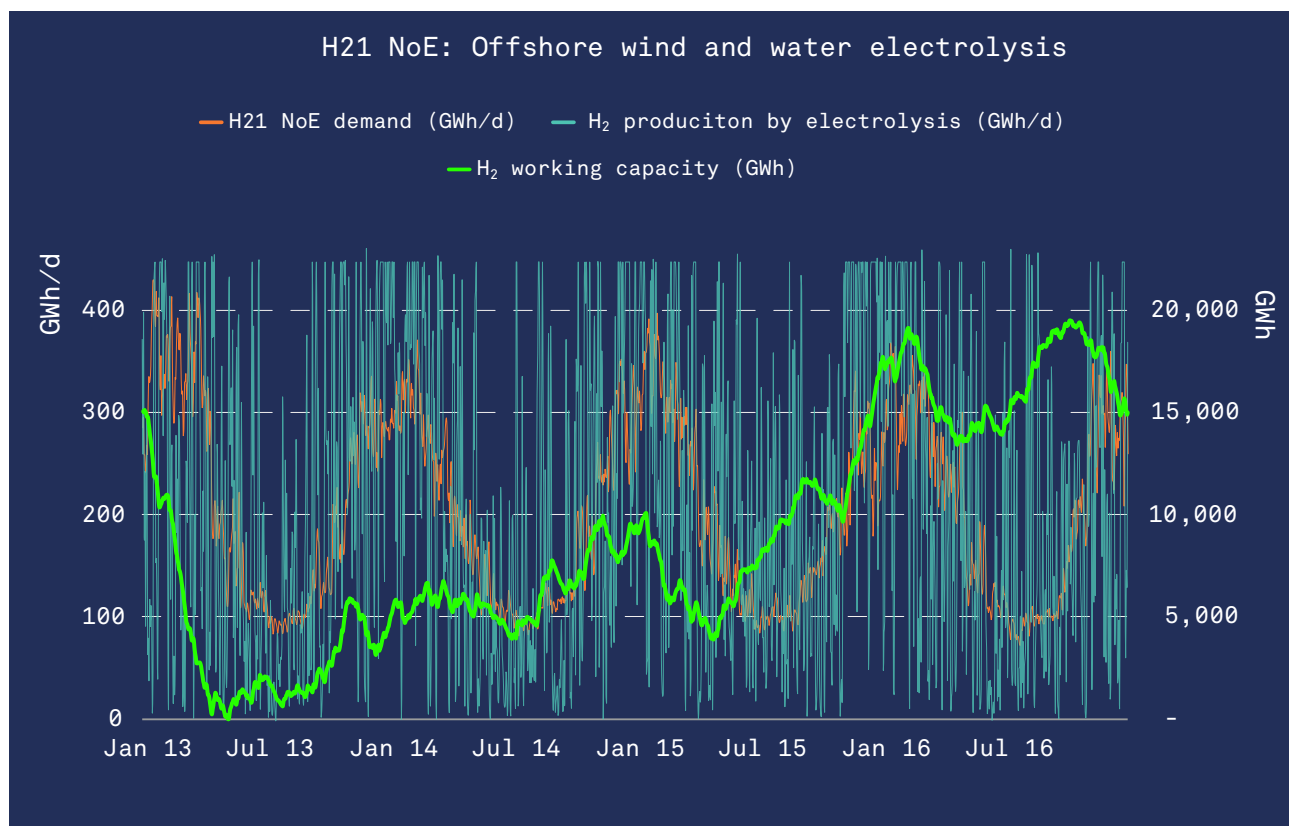


Figure 3.51: H21 NoE heat supply by offshore wind and water electrolysis

³² Renewable UK, Offshore Wind Project Timelines

Based on the electrolyser 1.5GW facility concept design (see Section 3.3) an 18 GW facility was designed (see Figure 3.52).

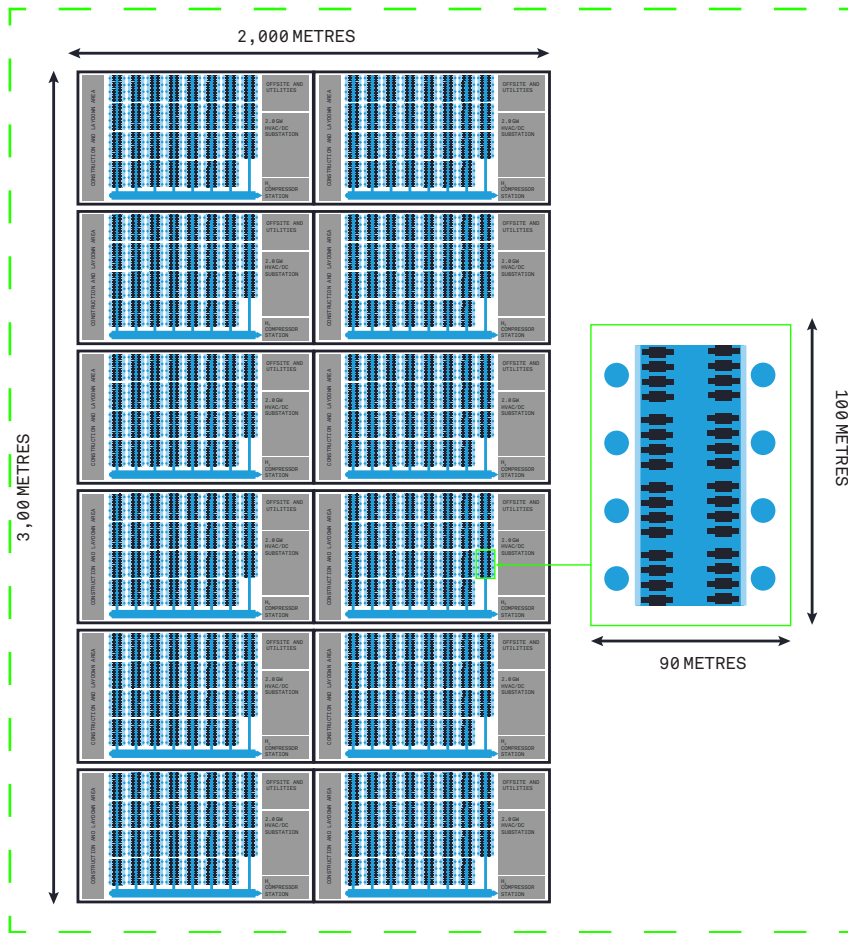


Figure 3.52: 18GW electrolyser based hydrogen facility for H21 NoE

The facility covers a vast area of 6,000,000 m² or 600 ha which would occupy a significant swage of land at any UK location. The facility, with possible location at Easington (see Section 4.0) is illustrated in Figure 3.53.

Based on the key performance data presented in [Section 3.3](#) it is possible to establish a similar dataset for the 18.4 GW hydrogen facility in addition to the CAPEX estimate.

It should be noted that the cost for the substation is slightly reduced since electricity from the wind is expected to be delivered as HVDC. This would result in 20% CAPEX reduction of the utility, offsites etc. and the specific cost will be reduced from £1,160/kWh_{H2HHV} to £1,073/kWh_{H2HHV}.

HYDROGEN PRODUCTION	MW _{H2HHV}	18,400
AVERAGE POWER CONSUMPTION	MW	25,952
ANNUAL WATER CONSUMPTION	m ³	18,900,000
COOLING WATER	m ³ /hr	61,300

Table 3.36: Key design parameters for a hydrogen facility based on water electrolysis

The total system CAPEX excluding the hydrogen transmission system (see [Section 5.0](#)) is shown in [Table 3.37](#).

AREA	CAPACITY	UNIT COST	TOTAL COST (£M)
Wind farm incl. HVDC platforms	26 GW	2,250 £/kW ¹⁾	58,500
HVDC transmission	26 GW	382.7 £/kW ²⁾	9,950
Hydrogen facility incl. substation	18.4 GW	1,073 £/kW _{H2HHV}	19,743
Hydrogen storage	19.5 TWh	225.7 £/MWh _{H2HHV}	4,401
TOTAL			92,595

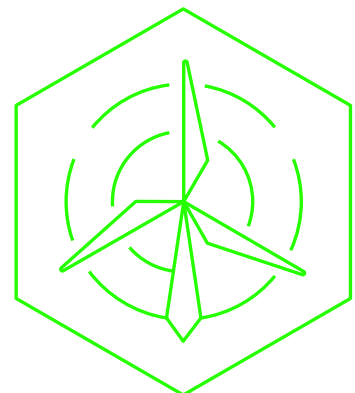
Table 3.37: Total system cost for an offshore wind via electrolysis to hydrogen supply concept

- **1)** From BEIS report electricity generation cost 2016. Offshore R3 – medium: pre-development: £120/kW and construction: £2,100/kW
- **2)** From BEIS report electricity generation cost 2016. Offshore R3 – medium: Infrastructure: £323m for 844 MW = £382.7/kW



Figure 3.53: Illustration of a H21 NoE 18GW electrolyser facility located at Easington

Costs withstanding, it is not considered practical to source low carbon hydrogen produced from offshore wind and electrolyser technology at the scale to convert the H21 NoE gas network and decarbonise heat by the later 2020s early 230s. Furthermore, the concept would require massive development of salt caverns, (i.e. 130), which is not practically deliverable within the time frame of the H21 NoE project ambitions or the UK governments CO₂ reduction obligations. The technical, project execution and commercial risks are simply too high.



3.9.2 Assessment of hydrogen from natural gas for the H21 NoE

Section 3.6 concluded that ATR is the preferred technology for the reforming of natural gas to hydrogen when high energy efficiency, high CO₂ capture rate and hydrogen at high pressure is required. A system design for H21 NoE was developed based on this resulting in a 12.15GW hydrogen plant comprising eight 1.5GW ATR units as the preferred option.

This would be coupled to inter-seasonal hydrogen storage using salt caverns, a hydrogen transmission system and CO₂ transport and storage (via pipeline) to an offshore aquifer in the North Sea for permanent storage. Further detailing on the system design is given in Section 4.0, 5.0, 6.0 and 7.0. A value chain with key figures is shown in Figure 3.54.

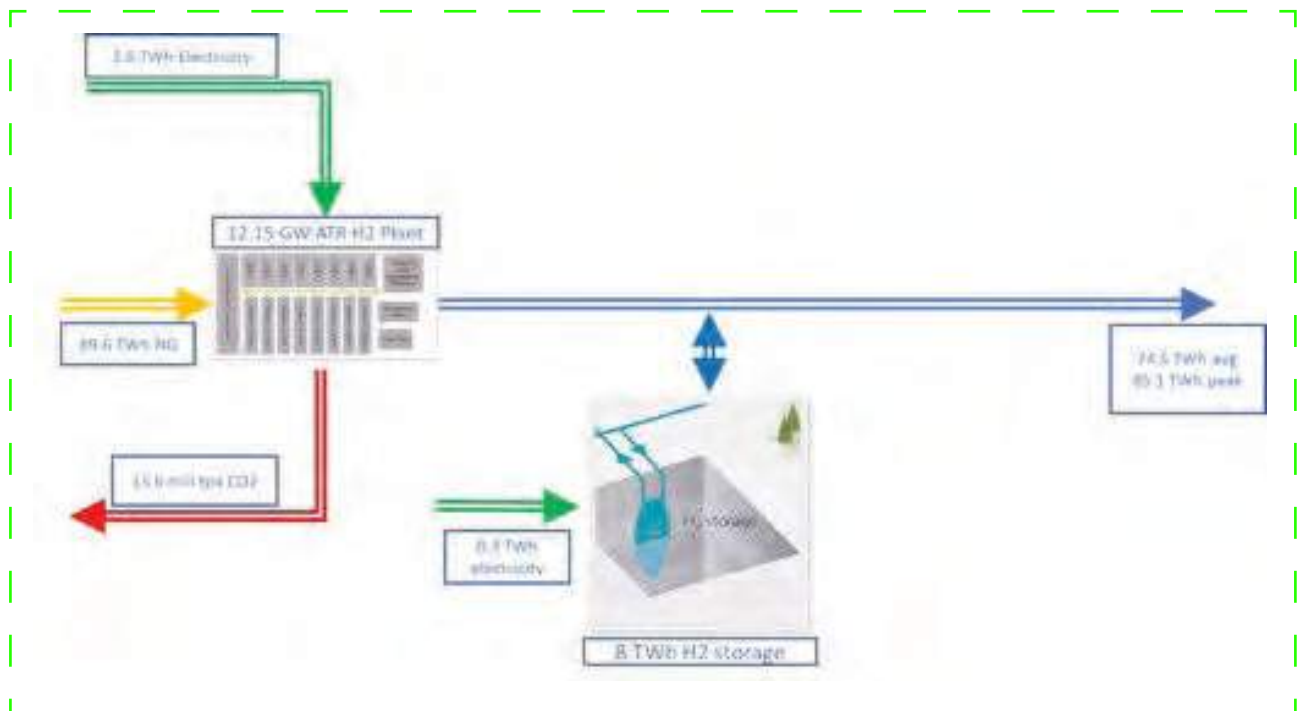


Figure 3.54: Value chain for the H21 NoE based on hydrogen from natural gas via ATR with CCS

The facility would need 14.6GW of natural gas and 588 MW of electrical import (equal to $0.048 \text{ kWh power/kWh}_{\text{H2HHV}}$).

The inter-seasonal hydrogen storage needed is circa 8TWh and the CO_2 storage would be 15-17 Mtpa.

As with electrolysis, the net emissions of the systems producing the hydrogen need to be considered to ensure such a strategy meets UK climate change obligations. For hydrogen produced via natural gas using either SMR or ATR technology, the carbon must be captured and stored long term to achieve the overall low carbon objectives. The CO_2 emissions from the hydrogen facility of the solution above are $13.1 \text{ g/kWh}_{\text{H2HHV}}$. In addition, there will be indirect emissions associated with the electricity import. For a grid emission of 100 g/kWh and 50 g/kWh the total direct and indirect emissions will be: $13.1 \text{ g/kWh}_{\text{H2HHV}}$ (ATR carbon footprint with imported power) + $(0.048 \text{ kWh/kWh}_{\text{H2HHV}} \times 100 \text{ g/kWh electricity import from grid}) = 17.9 \text{ g/kWh}_{\text{H2HHV}}$ and $13.1 \text{ g/kWh}_{\text{H2HHV}}$ + $0.048 \text{ kWh/kWh}_{\text{H2HHV}} \times 50 \text{ g/kWh} = 15.5 \text{ g/kWh}_{\text{H2HHV}}$.

In addition, supply chain emissions must be taken into consideration and initially this is considered to be delivered from Norwegian gas sources which have very low emissions (see [Section 9.0](#) for further details). Analysis of gas supply from other origins will be analysed and presented in [Section 9.0](#). However, the impact on the overall CO_2 footprint is shown in [Table 3.38](#).

NATURAL GAS ORIGIN	UPSTREAM EMISSION, $\text{G/KWH}_{\text{NGHHV}}$	H_2 PLANT EMISSIONS AND POWER IMPORT	TOTAL CO_2 FOOTPRINT, $\text{G/KWH}_{\text{H2HHV}}$
Norway Troll field	<1	15.5	16.0
UK average	26.3	15.5	47.1

Table 3.38: Impact of different natural gas sources on the overall CO_2 footprint of the natural gas value chain

As suggested above, this can only be surpassed or matched by water electrolysis if the electricity used comes from renewables or a combination of renewable and grid electricity which is not possible within the H21 NoE timescales. The availability of Carbon Capture and Storage (CCS) for permanent sequestration must also be considered. CCS is covered in detail in [Section 7.0](#) of this report. The main conclusion is that an ATR+CCS solution for H21 NoE is considered technically feasible within the required timeframe.

The estimated area needed for the facility would be approximately 90-120 ha or 15-20% of that required by the water electrolysis facility. Even though this is a mega scale hydrogen project this is not new to the industry. The Shell Pearl GTL project producing 140,000 bpd of synthetic fuels is based on oxygen blown ATR equal to 15GW of hydrogen production. In [Figure 3.55](#) is shown a Google map image of the Pearl GTL facility with highlighted areas of the air separation unit and ATR section.



Figure 3.55: Google Map image of Pearl GTL facility in Qatar

Commissioning

In contrast to water electrolysis the methane reforming technology market is mature and highly competitive, with decades of experience in serving a much larger market. In addition to 6 international ATR technology providers, there are numerous EPC contractors who have the experience in the engineering and construction of hydrogen facilities. A project the scale of H21 NoE is within the experience of the industry and the technical/project execution risk is low.

A commissioning strategy based on the ATR concept with inter-seasonal hydrogen storage in salt caverns securing the H21 NoE demand during the conversion period has been laid out in Section 6.0. The production facility has been designed to be a modular build adding one modular unit per annum from 2026 up to 2035. Excess hydrogen production in the early years can be managed via inter-seasonal storage, turn-down of individual units or the preferred solution of exporting as power to the UK electric grid. Inter-seasonal storage can be built incrementally in line with the conversion and commissioning strategy. However, the first four salt caverns would need to be in place by 2026. This puts the salt caverns on the critical path with respect to the project schedule.

The CO₂ transport and storage concept, to be discussed in detail in Section 7.0, would ideally be in place by 2026 for the H21 NoE to deliver clean hydrogen. It is concluded that both studied storage concepts on the UK and Norwegian continental shelves are feasible and that a CO₂ storage solution can be developed in time for a start-up of injection in 2026. This will require early involvement from UK Government with regards to access to acreage for offshore storage.

From a government perspective the H21 NoE project could be seen as a golden opportunity to achieve the CCS deployment levels needed to reach the 2050 target under the Climate Change Act.

A conversion of the H21 NoE heat supply to 100% clean hydrogen by an ATR based concept is considered feasible by 2035 provided that the FEED study starts in 2019 and that access to necessary acreage is given in time.

12.15GW H ₂ PRODUCTION	ATR OPTION 2	SMR OPTION 2
Carbon capture rate (%)	94.1	91.2
CO ₂ footprint (g CO ₂ /kwh)	13.1	20.5
Efficiency % (HHV)	79.9	79.5
CAPEX (£m) – Total	7,670	8,764
Electric Power Import (MW)	588	288.4
Natural Gas consumption (MW)	14,620	14,994
CO ₂ captured (t/hr)	2,552	2,527
Water consumption (m ³ per year)	5,600,000	11,200,000
Area (ha)	120-180	280-330
Configuration	8 ATR train + ASU	16 SMR trains

Table 3.39: Key design parameters for a hydrogen facility based on ATR and SMR

The total system CAPEX is shown for the ATR concept in [Table 3.40](#).

AREA	CAPACITY	UNIT COST	TOTAL COST (£M)
Hydrogen facility	12.15GW	£631/kW _{H₂HHV}	7,698
Hydrogen storage	8TWh	£225.7/MWh _{H₂HHV}	1,806
CO ₂ T&S	2,552 t/h	£460/t/h*	1,174
TOTAL			10,678

Table 3.40: Total system cost for an natural gas via ATR with CCS to hydrogen supply concept

* This is based on the Project Cost figures as set out in [Section 8.0](#).

3.9.3 Assessment of hydrogen by coal gasification for the H21 NoE

For completeness, hydrogen from coal gasification is analysed. This assumes coal is produced domestically in the UK. If coal import is used as a basis, it is important to include emissions related to bulk cargo transport.

For coal gasification similar design metrics as the ATR concept can be adopted, i.e. 12.15GW hydrogen production capacity and 8TWh inter-seasonal hydrogen storage.

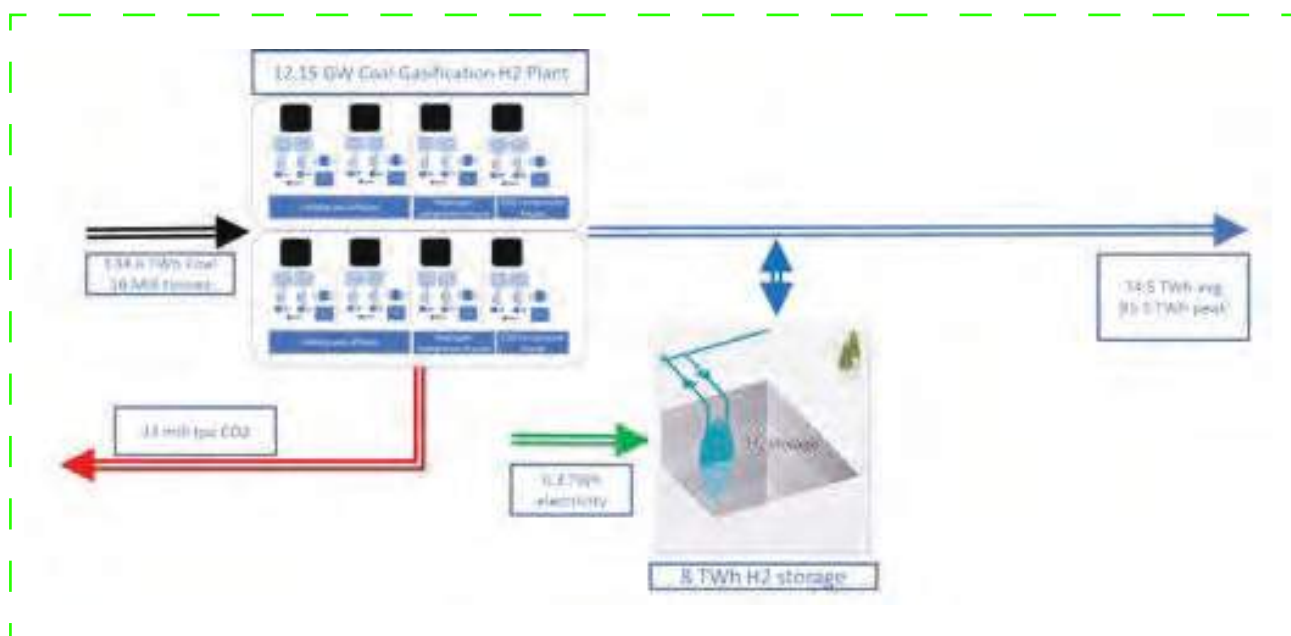


Figure 3.56: Value chain for the NoE based on hydrogen from coal gasification with CCS

Due to the higher carbon content in coal compared to natural gas, approximately twice as much CO₂ must be transported and stored (30-35 million tonnes per year). This would also require more CO₂ storage sites to be qualified resulting in increased cost and project complexity. This would likely result in CO₂ storage sites being put into operation post 2035. Therefore, the first CO₂ storage/sites would have to handle higher injection rates and reduce the potential total storage capacity (see Section 7.0).

As presented in Section 3.5, a coal gasification plant based on bituminous coal and 88% capture rate emits 60.3 g CO₂/kWh_{H2HHV}. An LCA assessment of UK based coal production³³ reported the GHG emissions from mining of coal was about 22.5 g CO₂/kWh_{CoalHHV} originating from methane leakage and energy consumption. Therefore, with an efficiency of 55.3% (HHV), the upstream emission is 41.2 g/kWh_{H2HHV} resulting in an overall CO₂ footprint of 101.5 g/kWh_{H2HHV}.

A hydrogen-based coal gasification process can result in only 50% CO₂ reduction compared to burning natural gas.

To produce 85.12TWh of hydrogen in a peak year, about 85.12/55.3%/8.44 kWh/kg = 18.2 Mt of coal will be consumed. During winter, when all coal gasification units are running, 12.15 x 24/55.3%/8.44 = 62,500 tonnes of coal will be consumed daily. Furthermore, 10,000 tonnes of ash/slag and sulphur will have to be transported from site. Therefore, it is most likely a UK based coal gasification plant would be co-located at existing coal power plants.

³³ N.A. Odeh, T.T. Cockerill "Life cycle analysis of UK coal fired power plants", Energy Conversion and Management 49 (2,008) 212-220

At the Secunda facility in South Africa, Sasol commissioned a 2 x 70,000 bpd coal to liquids facility equal to 14 GW of hydrogen with a unit entering operation in 1984 and another one in 1987. The site covers an area of 16,000,000 m² (4 x 4 km). Therefore a 12 GW hydrogen coal gasification facility is considered feasible to commission within the H21 NoE timeframe. About 5,000,000 to 8,000,000 m² of area would be needed for a 12.15 GW coal gasification to hydrogen facility.

The greatest challenge would be in the operation phase associated to environmental issues of handling huge amount of ash/slag, chemicals, sulphur and water.



Figure 3.57: Google maps image of Sasol 140,000 bpd coal to liquids facility at Secunda in South Africa

In order to mitigate some execution and operation risk, splitting the facility to multiple sites, e.g. 4-5 would be preferred. This would, however, result in a more complex and costly CO₂ transport infrastructure.

Based on the analysis of coal gasification in [Section 3.5](#) and 12.15GW coal gas gasification facility was established with key design parameters presented in [Table 3.41](#).

12.15GW H ₂ EXPORT	ENTRAINED FLOW GE GASIFICATION
Coal consumption (t/hr)	2,614
Coal consumption (MW _{HHV})	21,960
Efficiency (%-HHV)	55.3
CO ₂ -footprint (g/kWh)	60.3
CO ₂ capture rate (%)	88.0
Electric power consumption (MW)	1,780-2,025
CO ₂ export (t/h)	5,374
Water consumption (m ³ per year)	11,200,000
Area (ha)	500-800
Configuration	48 gasifiers at 457MW/h and 16 ASU units of 3,500 t/d

Table 3.41: Key design parameters for a hydrogen facility based on entrained flow GE gasification

The total system CAPEX is shown for the coal gasification concept with CCS in [Table 3.42](#).

AREA	CAPACITY	UNIT COST	TOTAL COST (£M)
Hydrogen facility	12.15GW	£1,691/kW _{H₂HHV}	20,546
Hydrogen storage	8 TWh	£225.7/MWh _{H₂HHV}	1,806
CO ₂ T&S	5,374 t/hr	£460/t/h	2,472
TOTAL			24,823

Table 3.42: Total system cost for an coal gasification with CCS to hydrogen supply concept

3.9.4 Assessment of hydrogen from ammonia for the H21 NoE area of conversion

Ammonia was discussed in [Section 3.7](#) in the context of intraday and inter-seasonal storage options.

The UK is well placed with the amount of natural gas it can produce so it is unlikely that large scale ammonia imports will be required in the short term.

However, the supply chain for a potential ammonia system is analysed below for completeness. An ammonia to hydrogen value chain based on natural gas is shown in [Figure 3.58](#) and has the following main components:

- Ammonia production in the US based on natural gas with CCS
- Ammonia shipped by large vessels to UK
- Ammonia cracked back into hydrogen in UK
- Inter-seasonal storage in UK

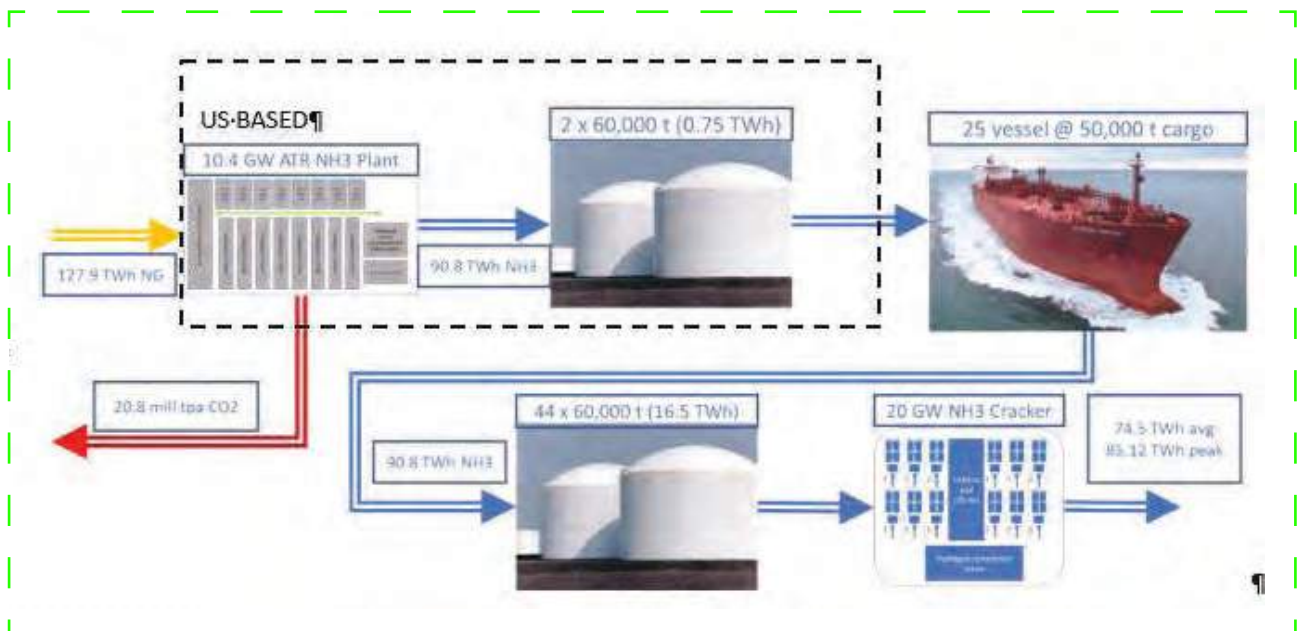


Figure 3.58: Ammonia value chain to the UK based on ammonia production in the US from natural gas with CCS

The carbon footprint of the ammonia plant can be compared to that of the ATR/SMR system producing the hydrogen for subsequent ammonia conversion. If this was undertaken in the UK, based on the system design outlined in [Section 4.0](#) of this report, the carbon footprint of the ammonia would be 20.6g/kWh. However, this rises to of 23 g/kWh when the additional energy penalty associated with the production and cracking process is added.

For H21 NoE the annual consumption of ammonia will be in the range of 13-14 Mtpa (about 10% of the current global production) and will require an ammonia production facility of close to 40,000 t/d. The biggest ammonia plants currently constructed have a capability of about 4,000 t/d. Therefore, a facility of 10 production units would be needed to supply the H21 NoE area of conversion with sufficient capacity.

Ammonia will be transported in VLGC (Very Large Gas Carriers) in liquid form at atmospheric pressure (-33°C). A typical VLGC has a capacity of 52,000 tonnes of ammonia (circa 325 MWh). This means that one VLGC must dock and load 52,000 tonnes of ammonia every day. The dock must accommodate two VLGC with simultaneous loading.

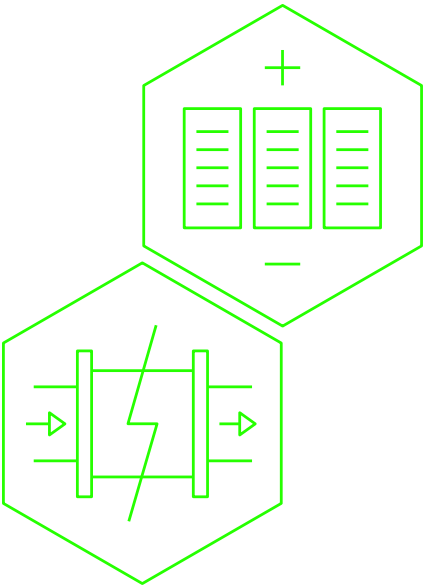
Based on the fuel and emission calculator by National Technical University of Athens – Laboratory for Maritime Transport³⁴, the fuel consumption and associated emission for a round trip shipping of ammonia from US Gulf Corpus Christi to UK Liverpool of 4,843 nautical miles³⁵ is shown in [Table 3.43](#).

KEY DATA	SINGULAR	H21 NOE
Ship type	LPG carrier	25 vessels
Size, DWT	52,000tonnes (325,000MWh)	13,400,000 tonnes ammonia annual equal 83,75TWh
Engine type	Slow speed	
Total distance	9,686nm	Total annual round trips: 258
Fuel consumption	1,340 tonnes equal 16.5GWh or 5% of the ammonia energy	345,300 tonnes or 4.25TWh annually
CO ₂ emission	4,167tonnes equal 12.8g/kWh	1.07Mt CO ₂ annual emissions

Table 3.43: Ammonia shipping performance data

Ammonia would be stored both at the production location and at the receiving facility in the UK. A typical ammonia tank can store about 60,000 tonnes (375,000 MWh).

A US based ammonia plant operating in base load, exporting at a constant rate and meeting the NoE demand would require 23 ship transports per month carrying 50,000 tonnes of ammonia each. The system would include 44 storage tanks in the UK (each storing 60,000 tonnes) in order to meet the peak month demand of 20 GW of hydrogen. Two such tanks would also be required in the US.



34 <http://www.martrans.org/emis/>
35 <https://sea-distances.org/>

The ammonia cracker capacity in the UK will also have to be designed to meet the 20 GW demand. An ammonia cracker unit will typically have a capacity of 400 MW requiring 50 units to meet the demand.

	ATR BASED AMMONIA	17 BAR AMMONIA CRACKING
Production capacity (MW)	10,400	20,000
Carbon capture rate (%)	92.1	0
CO ₂ footprint (g CO ₂ /kWh)	20.6	0
Efficiency % (HHV)	71.0	88.9
Electric power import (MW)	0	17.3
Natural gas consumption (MW)	14,650	N/A
Ammonia consumption (MW)	N/A	22,480
CO ₂ captured (t/hr)	2,384	N/A
Water consumption (m ³ per year)	7,000,000	0
Area (ha)	120-180	280-330
Configuration	8 ATR train + ASU	

Table 3.44: Key performance data for a natural gas to ammonia via ATR with CCS to hydrogen supply concept

Based on the above, the annual supply chain CO₂ emissions are as follows:

Ammonia plant: 1.87Mt (20.6g/kWh x 90.8TWh/y)

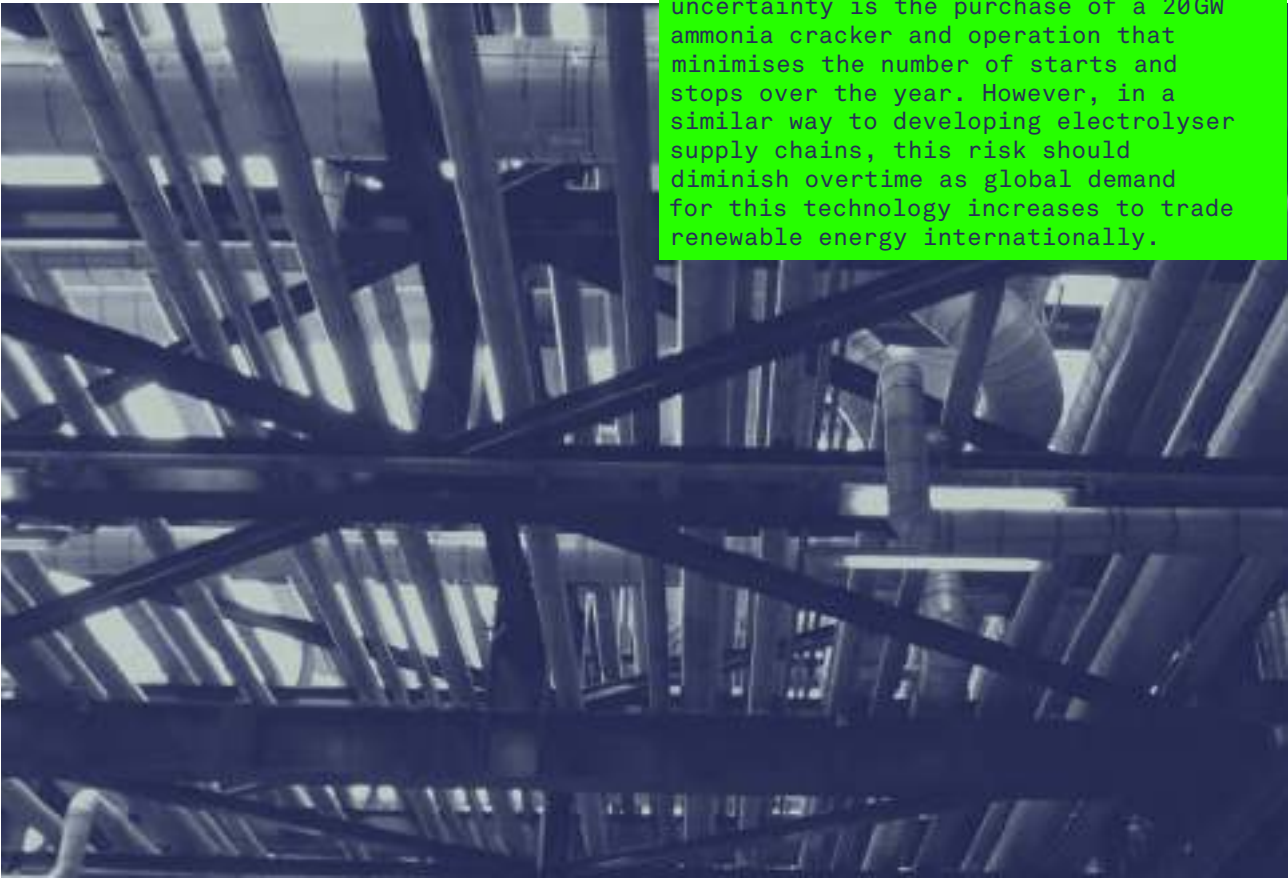
Transport: 1.07Mt

Total: 2.94Mt = 39.4g/kWh

The total system CAPEX is shown for the ATR based ammonia to hydrogen concept in [Table 3.45](#).

AREA	CAPACITY	UNIT COST	TOTAL COST (£M)
Ammonia production	10.4GW	£1,205/kW _{NH3HHV}	12,532
Ammonia storage and dock export facility	2 x 60,000 tonnes	£800/tonnes £100m per dock	296
CO ₂ T&S	2,384 t/hr	£460/t/h	1,097
Ammonia VLGC	25 x 50,000 tonnes	£40m per vessel	1,000
Ammonia storage and dock Import facility	44 x 60,000 tonnes	£800/tonnes £100m per dock	2,311
Ammonia cracker 17bar	20GW	£235/kW _{H2HHV}	4,700
TOTAL			21,937

Table 3.45: Total system cost for natural gas to ammonia via ATR with CCS to hydrogen supply concept



This concept is achievable within the time frame but the biggest risk and uncertainty is the purchase of a 20GW ammonia cracker and operation that minimises the number of starts and stops over the year. However, in a similar way to developing electrolyser supply chains, this risk should diminish overtime as global demand for this technology increases to trade renewable energy internationally.

3.9.5 Ammonia based on renewables

Ammonia can also be produced from renewable sources via electrolysis ('green ammonia'). Low cost storage and transport of ammonia offers an opportunity to utilise availability of cheap renewable sources, e.g. solar power in hot climates. The advantage of solar energy as a renewable source are as follows:

- Solar produces Direct Current (DC) power which is the requirement for electrolysis (no AC-DC converter is necessary)
- Solar in hot climates provides stable and continuous power compared to wind which is, by its nature, an intermittent power source. Obviously, there is a power production swing during the day, i.e. between daylight and hours of darkness at night. This is where Li-Ion batteries could act as a buffer between the solar power plant and the electrolyser to level the power production. This would allow for the electrolyser to operate continuously and make hydrogen/ammonia production more economic.

A generic solar to ammonia concept has been developed based on ideal solar conditions with a net capacity factor of 29%. To produce the same daily output rate as the ATR based ammonia concept of 10.4 GW x 24 = 249 GWh/d, the solar to ammonia facility would require the following:

- 57 GW Solar farm producing annually 147 TWh of power. The solar panel would cover an area of 500-600 km². For a UK location the net capacity factor would be approx 10-11% and would require about 3 times more installed solar capacity (150-170 GW).
- 179 GWh Li-battery to enable ammonia production during night at 50% turndown ratio
- 13.9 GW_{NH₃HV} alkaline electrolyser-based ammonia facility producing annually 90.8 TWh of ammonia

Figure 3.59 shows the main technologies in a solar to ammonia concept, the hourly production of solar power, power from battery, solar and ammonia production.

Downstream of the ammonia plant, the remaining parts of the value chain are similar to the ATR based scheme.



Figure 3.59: Ammonia value chain to the UK based on ammonia production in the US with solar power and via electrolysis

CO₂ emission only occurs in transporting ammonia and is 1.07 million tonnes per year resulting in an overall CO₂ footprint of $1.07 \div 74.45$ (H21 NoE heat demand) = 14.4 g/kWh. This is based on the CO₂ footprint of the vessels within the timescales for H21 NoE; over time these will reduce as shipping decarbonises. The total system CAPEX is shown for the solar to ammonia to hydrogen concept in Table 3.46.

AREA	CAPACITY	UNIT COST	TOTAL COST (£M)
Solar	57GW	£670/kW ¹⁾	38,190
Li-Battery	179GWh	£150/kWh ²⁾	26,850
Ammonia production	13.9GW	£1,594/kW _{NH₃HHV} ³⁾	22,157
Ammonia storage and dock export facility	2 x 60,000 tonnes	£800/tonnes £100Mt per dock	296
Ammonia VLGC	25 x 50,000 tonnes	£40m per vessel	1,000
Ammonia storage and dock import facility	44 x 60,000 tonnes	£800 tonnes 100Mt per dock	2,311
Ammonia cracker 17bar	20GW	£235 kW _{H₂HHV}	4,700
TOTAL			95,504

Table 3.46: Total system cost for solar power to ammonia via electrolysis to hydrogen supply concept

- **1)** From BEIS report electricity generation cost 2016. PV >5MW – medium: predevelopment: £70/kW and construction: £600/kW
- **2)** An optimistic cost <\$200/kWh have been used. For the note Lazard reports (LCOS 3.0) cost to be in the range of \$500/kWh with a potential to be reduced to \$300/kWh³⁶
- **3)** Reduced from £1,902/kW since NH₃ storage and dock is outside scope and substation simplification due solar – battery DC power production

If the electrolyser and ammonia plant could stop overnight, the lithium battery storage could be omitted and the plant would follow the solar production profile. This would increase the electrolyser/ammonia plant by a factor of 3.3 and would not create any savings. Furthermore, the potential for future cost reduction is more likely to occur for lithium batteries than electrolyzers.

For the H21 NoE timeframe, the installation of 57 GW of solar capacity (30% more than current entire German solar capacity) is considered very challenging.

An option based on importing green ammonia produced from solar will require significant capital and must be considered a longer-term option, i.e. post 2040.

³⁶ Lazard, "Lazard's Levelized cost of storage analysis – Version 3.0", 2,017

3.9.6 Ammonia based on gasification

Ammonia can be produced as part of coal gasification. This is a specific option being considered by the Carbonnet Project run by the Victoria Government in Australia. This may represent a credible source of internationally traded 'clean' hydrogen in the longer term for the UK. However, it is not considered practical in the context of the H21 NOE project as much more efficient, cheaper and cleaner systems are available directly to the UK via natural gas.

As with the ATR and electrolyser-based import ammonia to hydrogen supply concept, the ammonia production plant will deliver a base load production of $10.4 \text{ GW}_{\text{NH}_3\text{HHV}}$ and 90.8 TWh annually. This feeds into the transport, storage and cracker chain in both the ATR and electrolyser concept (see Figure 3.60). The supply chain will consume 184.5 TWh of coal equivalent to 22 Mt .

This is twice as much as the UK based Drax power station when operating on coal. This also results in an overall efficiency less than 40% ($74.5 \text{ TWh} / 188.75 \text{ TWh}$ (coal: 184.5 TWh + bunker fuel: 4.25 TWh) = 39.5% (HHV)).

45.2 Mtpa of CO_2 (3 times more compared to the ATR+CCS based hydrogen concept) would need to be captured and must be stored. For a 40 year operation, this would require $1,800 \text{ Mt}$ of CO_2 storage capacity.

The overall CO_2 emission would be:

Ammonia plant: 5.9 Mt
 $(65.1 \text{ g/kWh}_{\text{NH}_3\text{HHV}} \times 90.8 \text{ TWh})$

Transport: 1.1 Mt

Total: 7.0 Mt or $94 \text{ g/kWh}_{\text{NH}_3\text{HHV}}$

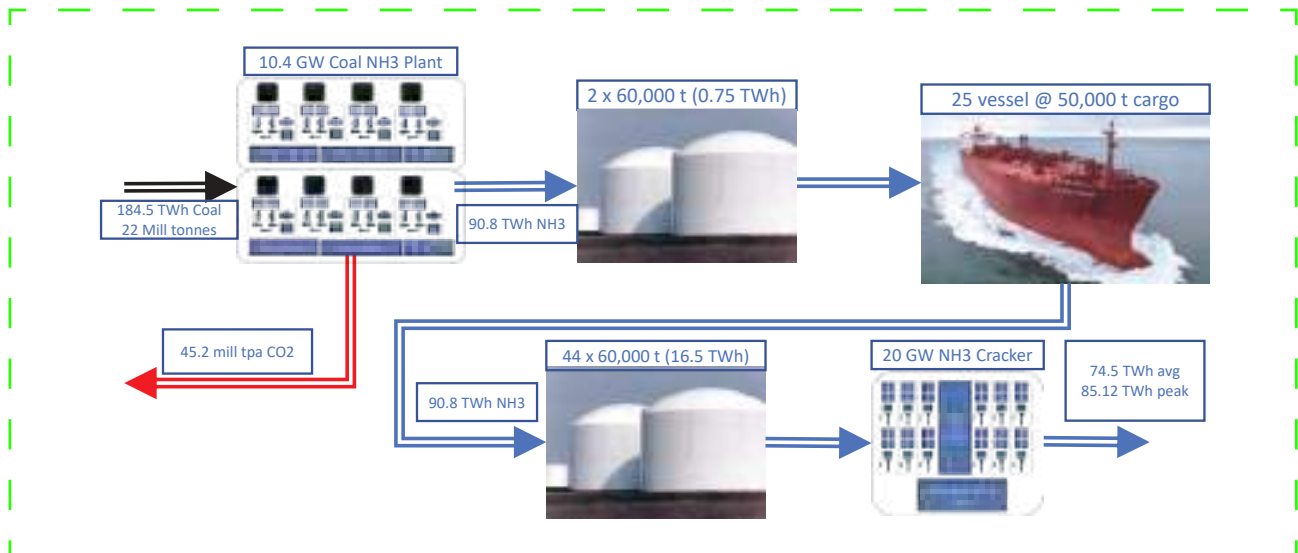


Figure 3.60: Ammonia value chain to the UK based on ammonia production in the US from coal gasification with CCS

The total system CAPEX is shown for the ATR coal gasification to ammonia concept in [Table 3.47](#).

AREA	CAPACITY	UNIT COST	TOTAL COST, £M
Ammonia production	10.4GW	£2,350/kW _{NH3HHV}	24,440
Ammonia storage and dock export facility	2 x 60,000 tonnes	£800/tonnes	296
CO ₂ T&S	5,159 t/hr	£460/t/h	2,373
Ammonia VLGC	25 x 50,000 tonnes	£40 million per vessel	1,000
Ammonia storage and dock Import facility	44 x 60,000 tonnes	£800/tonnes	2,311
Ammonia cracker 17bar	20GW	£235/kW _{H2HHV}	4,700
TOTAL			35,121

Table 3.47: Total system cost for a coal to ammonia via gasification with CCS to hydrogen supply concept



3.10

Conclusions and final technology selection

A qualitative comparison of the different value chains considering essential selection criteria for the H21 NoE conversion is given in [Table 3.48](#).

ENERGY SOURCE SUPPLY CHAIN SOLUTION	UNIT	NATURAL GAS ATR W CCS	OFFSHORE WIND AEL	COAL GASIFICATION W CCS	NATURAL GAS ATR W CCS NH ₃ CRACKING	COAL GASIFICATION W CCS NH ₃ CRACKING	SOLAR AEL NH ₃ NH ₃ CRACKING
ANNUAL DEMAND	TWh	74.5	74.5	74.5	74.5	74.5	74.5
KEY PERFORMANCE PARAMETERS							
FOSSIL FUEL CONSUMPTION	TWh	89.6	0	134.60	132.15	188.75	4.25
ELECTRICITY CONSUMPTION	TWh	3.9	105.9	0.30	0	0	147
TOTAL	TWh	93.5	105.9	134.90	132.15	188.75	151.25
OVERALL SUPPLY CHAIN EFFICIENCY	%	79.7%	70.3%	55.2%	56.4%	39.5%	49.3%
CO ₂ CONSIDERATION							
CO ₂ FOOTPRINT EXCL SUPPLY CHAIN EMISSIONS	g/kWh	15.5	0.0	60.3	39.4	94.0	14.4
CO ₂ FOOTPRINT INCL SUPPLY CHAIN EMISSIONS	g/kWh	16.5-50	0.0	101.5	40-65	139.6	14.4
CO ₂ CAPTURED	Mtpa	15.6	0.0	33.0	20.8	45.2	0.0
DESIGN CAPACITIES							
RENEWABLE ELECTRICITY PRODUCTION	GW		26.0				57.0
RENEWABLE ELECTRICITY TRANSPORT	GW		26.0				57.0
RENEWABLE ENERGY STORAGE	GWh						179.0
AMMONIA PRODUCTION	GW				10.4	10.4	13.9
AMMONIA TRANSPORT AND STORAGE	TWh				90.8	90.8	90.8
HYDROGEN PRODUCTION	GW	12.2	18.4	12.2	20.0	20.0	20.0
HYDROGEN STORAGE	TWh	8.0	19.5	8.0	0.0	0.0	0.0
CO ₂ T&S	t/hr	2,552.0		5,374.0	2,384.0	5,159.0	0.0
CAPEX							
RENEWABLE ELECTRICITY PRODUCTION	£m		58,500				38,190
RENEWABLE ELECTRICITY TRANSPORT	£m		9,950				
RENEWABLE ENERGY STORAGE	£m						26,850
AMMONIA PRODUCTION	£m				12,532	24,440	22,157
AMMONIA TRANSPORT AND STORAGE	£m				36,07	3,607	3,607
HYDROGEN PRODUCTION	£m	7,698	19,743	20,546	4,700	4,700	4,700
HYDROGEN STORAGE	£m	1,806	4,401	1,806			
CO ₂ T&S	£m	1,174		2,472	1,097	2,373	
TOTAL CAPEX	£m	10,678	92,595	24,823	21,936	35,121	95,504
ANNUAL OPEX							
FOSSIL ENERGY PRICE	£/MWh	23		9	12	5	
ELECTRICITY PRICE	£/MWh	60		60	30	30	
CO ₂ EMISSION PRICE	£/t	80	80	80	80	80	80
FOSSIL ENERGY COST	£m	2,061	0	1,211	1,586	944	0
ELECTRICITY COST	£m	234	0	18	0	0	0
CO ₂ COST	£m	92	0	359	235	560	86
O&M (3 % OF CAPEX)	£m	320	2,778	745	658	1,054	2,865
TOTAL OPEX AND ENERGY	£m	2708	2778	2,333	2,479	2,558	2,951
H ₂ COST	£/MWH	50	155	63	61	79	160

Table 3.48: H21 NoE value chain summary

CO₂ reduction potential

To meet significant/deep CO₂ reduction in heat solutions a carbon capture rate of 85-90% is required.

With a starting point of 184.45 g/kWh (or 206 g/kWh including supply chain emissions) an 85-90% CO₂ reduction would mean that a concept should have a maximum CO₂ footprint of 15-30 g/kWh. As seen from the table only electrolysis directly connected to a renewable source (or at least 70-75% depending on the grid CO₂ footprint) and natural gas-based solutions can meet this target. Coal concepts whether it is hydrogen or ammonia have CO₂ footprint of 60-140 g/kWh.

Fossil fuel availability

For fossil-based supply the long-term availability is critical. An investment in H21 NoE would have a lifetime of 40 years (last plant would be in operation in 2070). Coal resources are immense with 400 years of reserves based on current annual fossil-based consumption. The UK in 2017 consumed 80 bcm of natural gas (800 TWh) of the global consumption of 3,670 bcm (36,700 TWh). Total reserves are 193,000 bcm and in the last 20 years 70,000 bcm of new reserves have been discovered which is equal to the consumption³⁷ over the same period.

Commissioning risk

The most important question is: Is it possible to commission the supply chain within the timeframe and conversion strategy of H21 NoE?

Looking at the different technology elements in the supply chain and estimating annual installation to meet H21 NoE targets, the key findings can be summarised as follows:

- Natural gas reforming by SMR/ATR is already being used on global scale with installations equal to 6-7 GW per year. References like Oryx and Pearl GTL projects are examples of mega scale projects similar or above H21 NoE requirements.
- Coal gasification is a technology that can be adopted for H21 NoE and the Sasol CTL facility in Secunda, South Africa is a project similar to H21 NoE.
- Ammonia production capacity, including ammonia storage and transport systems, would be possible to establish. 90.8 TWh capacity is less than 10% of current global production of ammonia and well within industry experience and scale.
- For all concepts with electrolysis about 1.5-2.5 GW_{H₂HHV} AEL capacity would have to be installed per year. This is 4-5 times greater than current big scale manufacturers can deliver in a 5-year time frame.
- 26 GW of offshore wind capacity would have to be commissioned for the H21 NoE green hydrogen supply concept. A typical mega scale project in the industry is about 1.2-2.4 GW. Delivering a 10-15 times larger scale project would be an enormous challenge within the H21 NoE project scope. This would also be the case for commissioning 57 GW solar power capacity.

³⁷ BP Statistical review of world energy, June 2018

ENERGY SOURCE SUPPLY CHAIN SOLUTION	UNIT	NATURAL GAS ATR W CCS	OFFSHORE WIND AEL	COAL GASIFICATION W CCS	NATURAL GAS ATR W CCS NH ₃ CRACKING	COAL GASIFICATION W CCS NH ₃ CRACKING	SOLAR AEL NH ₃ NH ₃ CRACKING
DESIGN CAPACITIES							
RENEWABLE ELECTRICITY PRODUCTION	GW		26.0				57.0
RENEWABLE ELECTRICITY TRANSPORT	GW		26.0				57.0
RENEWABLE ENERGY STORAGE	GWh						179.0
AMMONIA PRODUCTION	GW				10.4	10.4	13.9
AMMONIA TRANSPORT AND STORAGE	TWh				90.8	90.8	90.8
HYDROGEN PRODUCTION	GW	12.2	18.4	12.2	20.0	20.0	20.0
HYDROGEN STORAGE	TWh	8.0	19.5	8.0	0.0	0.0	0.0
CO ₂ T&S	t/hr	2,552.0		5,374.0	2,384.0	5,159.0	0.0
CO ₂ STORAGE CAPACITY	£m	600		1,200	750	1,800	
COMMISSIONING REQUIREMENT							
RENEWABLE ELECTRICITY PRODUCTION			2.5-3.0				7-7.5
RENEWABLE ELECTRICITY TRANSPORT			2.5-3.0				7-7.5
RENEWABLE ENERGY STORAGE							20-25
AMMONIA PRODUCTION					1.3	1.3	1.7
AMMONIA TRANSPORT AND STORAGE					4-5	4-5	4-5
HYDROGEN PRODUCTION		1.5-1.8	2.3-2.8	1.5-1.8	2.5	2.5	2.5
HYDROGEN STORAGE		6-7	15-17	6-7			
CO ₂ T&S		200KM PIPELINE 1-2 WELLS PA		200KM PIPELINE 2-4 WELLS PA	200KM PIPELINE 1-2 WELLS PA	200KM PIPELINE 2-4 WELLS PA	
CO ₂ STORAGE CAPACITY		QUALIFY 2-3 FORMATION		QUALIFY 4-6 FORMATION	QUALIFY 2-3 FORMATION	QUALIFY 6-9 FORMATION	

Table 3.49: Commissioning requirements for H21 NoE hydrogen supply concepts

