Techno-economic analysis of green hydrogen transport for long distance pathways

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Una pagina non potrà mai racchiudere ciò che questi anni hanno significato per me, quanto hanno influenzato la persona e, spero, l'uomo che sono adesso. Non solo pomeriggi di studio ed esami, molto di più: infiniti volti, legami indissolubili, luoghi magici, risate interminabili, pranzi tra amici la Domenica, ore di sonno perdute.

Oggi si chiude un cerchio o forse un circuito, perché di curve e irregolarità ne è stato pieno, ma non importa; rimarrà nei miei ricordi perfetto così come lo è adesso.

Un viaggio perché sia memorabile non può che essere condiviso con dei compagni e in questi anni in tanti si sono aggiunti lungo il tragitto, lo hanno abbandonato strada facendo o ne hanno sempre fatto parte. E queste righe che seguono sono per ringraziare tutti voi.

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Abstract

The first part of the thesis provides an insight within the Hydrogen (H_2) framework, starting from its main properties and safety-related issues and ending with an introduction to the current and future trends of the Hydrogen market.

The following chapters of the analysis (2,3,4) will put a focus on the main supply chain features of three ways of delivering H_2: Liquid Hydrogen (LH_2), Liquid Ammonia (NH_3) and Hydrogen Pipelines.
A small introduction regarding properties and market potential is present for each technology at the beginning of every chapter. Afterward, the main components of each system are analyzed and discussed from both an economic and technical point of view assigning, where possible, cost functions to estimate capital expenditure, and energy consumption functions to estimate operating expenditure.

Based on the cost´s projections coming from the previous analyses, the Levelized Cost Of Hydrogen (LCOH), main focus of this master thesis´ dissertation, is obtained. Since the LCOH is not a univocal outcome, but rather it depends on the specifics of every project, a sensitivity analysis accounting for the parameters that affect the LCOH the most is carried out at the end of every technology´s investigation.

Once a dynamic model has been developed for all the modalities of transport, a hypothetical scenario where Hydrogen is exported from a country with a forecasted low Renewable electricity´s cost (Casablanca, Morocco) to a potential future hub of Hydrogen (Marseille, France) is taken as a case study. The LCOH is again debated according to the specific input of the case study, highlighting weakness and point of strength of each technology and pointing out at the best design choice of such a system.

Finally, the conclusions gather all the master thesis´ highlights indicating for which range of production and distance of transport LH_2, NH_3 or Hydrogen pipelines are best suited for the purpose of delivering Hydrogen from one point to another.

Index

Nomenclature

Symbol	**Unit of measure**	**Description**
API	-	American Petroleum Institute
ASU	-	Air Separation Unit
BOG	%	Boil Of Gas
CAPEX	€	Capital Expenditure
CMS	-	Carbon molecular sieve
FCP	-	Fatigue crack propagation
GHG	-	Green House Gases
HAZ	-	Heat affected zone
HB	-	Haber-Bosch
H_2	-	Hydrogen
IDEALHY	-	Integrated Design for Efficient Advanced Liquefaction of Hydrogen
IRENA	-	International Renewable Energy Agency
LCOA	€/kgH_2	Levelized Cost Of Ammonia
LCOH	€/kgH_2	Levelized Cost Of Hydrogen
LGC	-	Large Gas Carriers
MGC	-	Medium Gas Carriers
NG	-	Natural Gas
NH_3	-	Ammonia
OPEX	€	Operative Expenditure
O&M	€	Operation and Maintenance
SMR	-	Steam Methane Reforming

Introduction

The level of penetration of Hydrogen in the future global energy framework is still unknown, but according to several government and private reports, it will be any way remarkable on the green transition of different sectors. Indeed, Hydrogen has the potential to become the second main energy carrier, after electricity, for the decarbonization of end-use sectors [1].

Hydrogen is a multipurpose energy vector: it can be used as a feedstock, a fuel, an energy carrier, and an energy storage medium, embracing many possible applications in the industry, transport, power and building sectors.

Even though the discussion around Hydrogen seems to have arisen recently, Hydrogen and energy have a long-shared history that goes back 1800s, when the first demonstration of water electrolysis and fuel cells has been performed. Thereafter, in the following centuries Hydrogen has provided lift to balloons since the 19th century, and propelled humanity to the moon in the sixties during the cold war.

In the most recent years, Hydrogen has been of fundamental importance in the synthesis of ammonia, which helped feeding a growing population. At the same time, H_2 has been an integral part of the oil refining industry since the mid-20th century [2].

The role of Hydrogen is now spreading according to new market demands, as transportation, storage and building heating. In a world always more surrounded by Renewables, complementing other storage applications, as hydro pumps and batteries, it can help keeping the electricity grid network reliable and safe. Furthermore, there is also a considerable potential for Hydrogen to repurpose the current Natural Gas (NG) infrastructure, avoiding additional pollution due to the construction of a new network [2].

The demand for hydrogen is around 70 million tons per year, but unfortunately this hydrogen is almost entirely supplied from fossil fuels, coal and NG on top. Consequently, the production of hydrogen is responsible annually for an emission of CO2 equivalent to the emission of Indonesia and the United Kingdom combined [3]. Therefore, the production of Hydrogen needs to quickly be switched to sustainable processes to avoid the release of considerable quantity of Green Houses Gases (GHG) in the next decades, against the goals for 2050.

Nowadays Hydrogen has appeal because is light, storable, reactive, has high energy content per unit of mass, it can be readily produced at industrial scale, and it does not produce any GHG during its combustion. But, as it happens for other energy carriers, Hydrogen is not free from health and safety risks when used on a large scale. Nevertheless, health and safety considerations of most hydrogen-based fuels and feedstocks are known and familiar to the energy sector, but the risks need to be well communicated and managed to users, especially in the early phase of deployment, as otherwise hydrogen may encounter some of the problems that have slowed down the deployment of the nuclear industry [3].

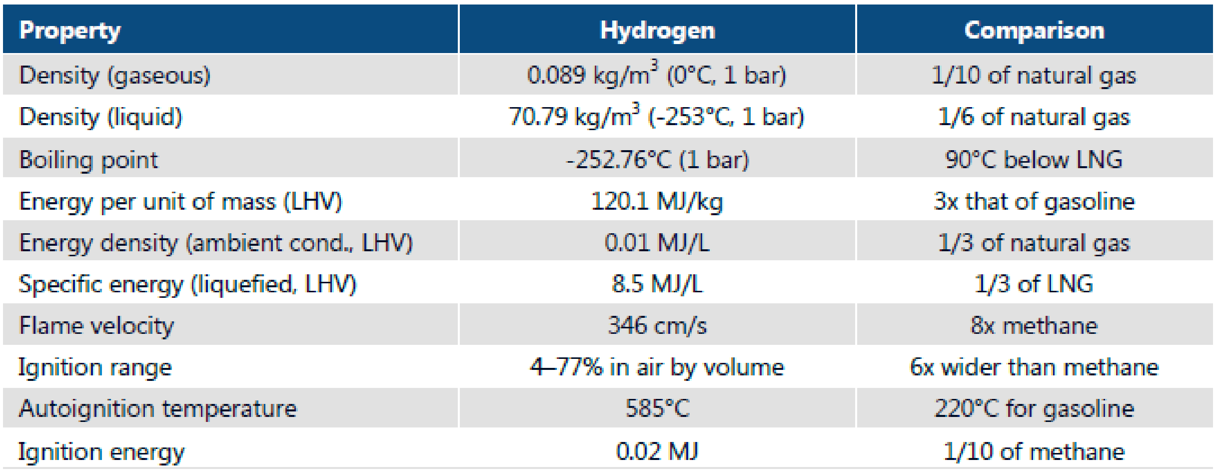
As a matter of fact, Hydrogen nowadays is still too expensive to compete with the traditional fossil fuels. To claim this position in the energy mix, Hydrogen will require further research and innovation into the production technologies and on the infrastructure network that can bring it to a geographically spread market [2].

1 Hydrogen

1.1 Physical properties

Hydrogen is a non-toxic, colorless and odorless gas, but it has a high flame velocity, broad ignition range and low ignition energy which make it a gas highly flammable and consequently dangerous to handle. These properties are in some ways partly counterbalanced by its high buoyancy and diffusivity, which, in an open environment, make it easy to dissipate quickly. Some of the main properties of Hydrogen are included in Table 1.

Table 1 - Hydrogen main properties – table from [3]



Property	Hydrogen	Comparison
Density (gaseous)	0.089 kg/m^3 (0°C, 1 bar)	1/10 of natural gas
Density (liquid)	70.79 kg/m^3 (-253°C, 1 bar)	1/6 of natural gas
Boiling point	-252.76°C (1 bar)	90°C below LNG
Energy per unit of mass (LHV)	120.1 MJ/kg	3x that of gasoline
Energy density (ambient cond., LHV)	0.01 MJ/L	1/3 of natural gas
Specific energy (liquefied, LHV)	8.5 MJ/L	1/3 of LNG
Flame velocity	346 cm/s	8x methane
Ignition range	4–77% in air by volume	6x wider than methane
Autoignition temperature	585°C	220°C for gasoline
Ignition energy	0.02 MJ	1/10 of methane

Hydrogen at its purest form, has a very low density (0,089 kg/m^3), which make it almost impossible to manage in any sector without compression or liquefaction treatments. Indeed, being Hydrogen the lightest element in the periodic table, it has a low energy density per unit of volume, meaning that larger volumes of hydrogen must be moved to match identical energy demands as compared with other fuels. Yet, Hydrogen has 3 times more the energy per kg (120 MJ/kg) than natural gas.

Always because of its light weight, particular attention should be paid to the tank or pipe in which is stored or is passing through. Indeed, Hydrogen is so small it can diffuse into some materials, even bonding with the elements, leading to cracks or failure. It also escapes more easily through sealings and connectors than larger molecules, such as natural gas, making necessary some precautions new for some industry [3].

The Boiling point of Hydrogen is extremely low, roughly -253°C; such a low temperature needs an enormous quantity of energy to be reached, making the Liquid Hydrogen one of the most expensive energy carrier to produce. Nevertheless, Hydrogen in its liquid form has a higher energy density that Hydrogen compressed at 750 bar, reason why in sectors where the space is of critical importance, liquid Hydrogen can be a meaningful energy carrier.

1.2 Market

The future global hydrogen market is strictly bounded on both the technological improvements and political willing. [4] states that around 350 projects of hydrogen production from electrolysis are currently under development, which could bring the global capacity up to 54 GW by 2030. Moreover, another 40 projects accounting for more than 35 GW of capacity are in early stages of development. At the current state, the market of Hydrogen is dominated by the global demand coming from the Ammonia production needed for urea and other fertilizers, as it is shown in Figure 1.

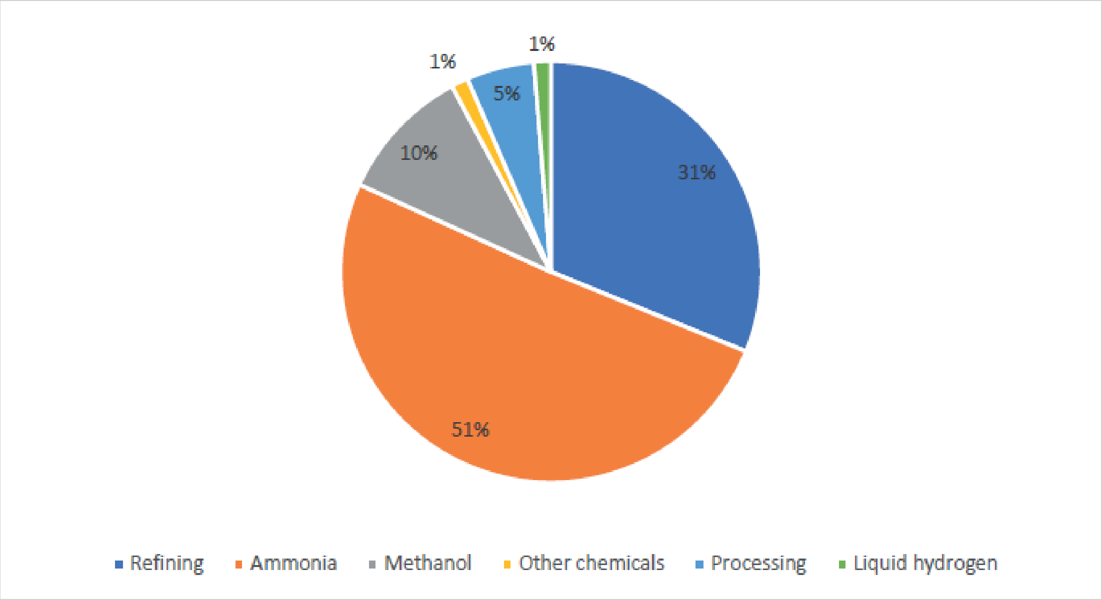


Figure 1 - Global hydrogen demand by subsector - image from [5]

Of the remaining half, Hydrogen is used for around 30 percent in various processes related to refineries and about ten percent is used for methanol production. According to one report or the other, the share of Hydrogen on the final energy mix can be moderate or very large. International Renewable Energy Agency (IRENA) estimate an additional market demand of 8 EJ in 2050 in addition to the current demand in feedstock, projecting the major growth in the transportation sector. On the other hand, Shell argues that hydrogen will emerge only from 2040 onwards, firstly for industry and transport, with a growth of about 8 EJ until 2050 and then with a steep increase onwards. In a more optimistic scenario, the Hydrogen Council forecasts a tenfold increase on the hydrogen market in 2050, with the transportation sector leading the rising trend. The European commission in its strategic vision for a climate-neutral EU projected the share of hydrogen in Europe’s energy mix to grow from the current less than 2% to 13-14% by 2050 [2]. Looking at Figure 1 Hydrogen for transporting purposes is merely a marginal market as of today, as it is the building heating. Yet, big efforts from the industries in the last decades, have make the cost of automotive fuel cells fallen by 70% since 2008 thanks to technological progress and growing sales of fuel cell electric vehicles (FCEVs) [4]

2 Transport of Liquid Hydrogen

2.1 Introduction

The liquefaction process of hydrogen it has been studied since the beginning of the 19th century, and it underwent a huge boost during the Fifties and Sixties because of the aerospace industry. During those years the prototype and small-scale plants completely switch their capacities to meet the demand of the space race [6].

The current global liquefaction capacity is around 350 tons/day and it has to match the demand coming from aerospace, chemical, electronic/semiconductor and metallurgical industries. Globally, the capacity is not spread evenly, because 300 out of 350 tons per day of liquid Hydrogen are produced in USA and Canada, mainly because of aerospace industry need. Once the demand will rise in the future, this capacity will not be enough anymore, and new plants are going to be built around the world.

When Hydrogen is cooled down to its liquid form, it increases the density from 0,089 kg/m^3 to 70,8 kg/m^3. This change in density can be beneficial for several applications where Hydrogen is currently put aside for the lack of space, embracing new sectors where even the compressed Hydrogen could not get into.

The supply chain of Hydrogen that will be analyzed in the next sections is shown in Figure 2.

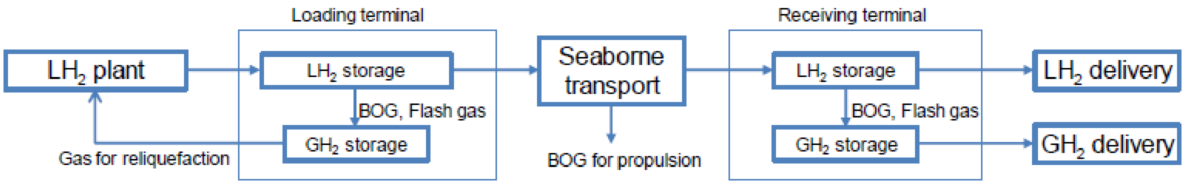


Figure 2 - Overall system of LH2 transport - image from [7]

From the liquefaction plant, Hydrogen is brought to the import or loading terminal where is temporarily stored waiting for being sent to the ship or sent directly to the ship´s tanker. The capacity of the terminal varies depending on the quantity delivered by the liquefaction plant, and also on the frequency with which the ships come to the terminal to load and unload Hydrogen. Longer period of storage implies bigger terminals. The frequency of the shipment is also function of the number of the ships, the velocity of the ships and the distance of transport.

Once the ship, is totally refueled with LH_2, it leaves the terminal towards the unloading terminal, where the Hydrogen will be again stored or directly delivered. Depending on the size of the terminal, both the liquified and gaseous form of Hydrogen can be sold.

Unluckily, besides the storage already being used in the aerospace industry, most of the supply chains of liquid Hydrogen is missing or is at the prototype level. This has implications on the availability of data needed to define the cost of the system. In the few works found, an economy of scale approach is often utilized. This kind of approach that will be defined afterwards in the next sections could lead to underestimate the costs, since is usually applicable for established sectors and not emerging ones where the uncertainties are still too much. Therefore, when the literature suggests adopting this method for the cost evaluation of a system component, this analysis will present an alternative approach in parallel.

2.2 Physical properties

Liquid hydrogen is odorless, tasteless, and colorless. Because of its properties and the extreme conditions required to produce it, handling liquid hydrogen represents a technical challenge for the industry. Indeed, its boiling temperature is just 20°C above the absolute zero, meaning that it needs enormous quantities of energy to be liquefied. One of the main differences between gaseous hydrogen and liquid is the value of density. Indeed, in many applications, like the automotive industry, the space required by the confinement of the compressed hydrogen is not a secondary problem, lowering the pace of the transition. Other important properties are listed in Table 2 below.

Table 2 - Liquid Hydrogen main properties

Property	Value	Unit
Liquid density (@-253°C, 101.3kPa)	70,86	kg/m3
Critical pressure	1,3	MPa
Critical temperature	-240	°C
Boiling point	-253	°C
Heat of vaporization	461	kJ/kg
Lower Heating Value (LHV)	118,8	MJ/kg
Heat capacity (@-256°C)	8,1	kJ/(kg*°C)

Molecular hydrogen occurs in two isomeric forms, one with its two proton nuclear spins aligned parallel, that takes the name of ortho-hydrogen, the other with its two proton spins aligned antiparallel, called para-hydrogen. Depending on the temperature of the environment where hydrogen lies and the heat provided, the equilibrium between these two isomers changes. For example, at room temperature and thermal equilibrium, thermal excitation causes hydrogen to consist of approximately 75% ortho-hydrogen and 25% para-hydrogen. Figure 3 shows how the concentration of para-hydrogen increases significantly for low temperature, under equilibrium conditions.

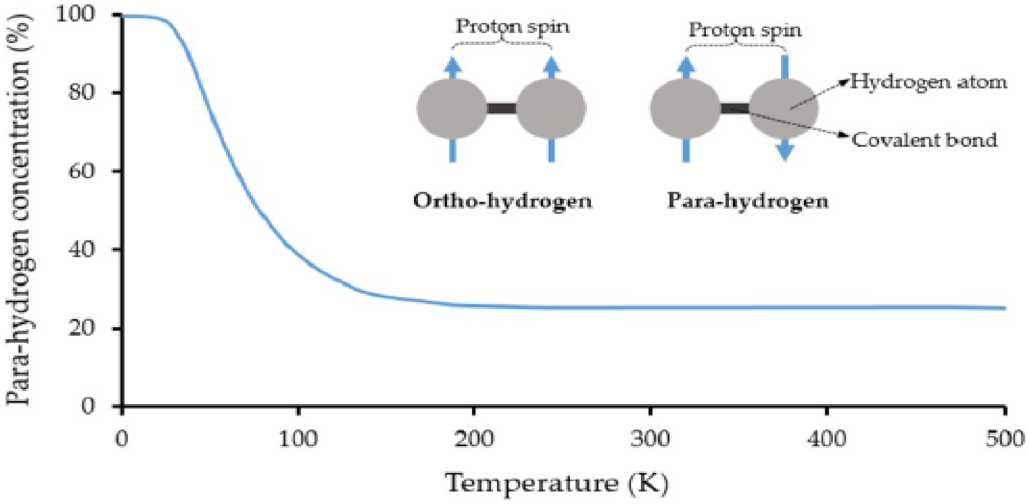


Figure 3 – Ortho-Para Hydrogen equilibrium at different temperature - image from [6]

There is no significant change in the concentration ratio of both ortho- and para-hydrogen at temperatures higher than −120 ◦C (150 K). However, when the temperature is cooled to near zero, the concentration of para-hydrogen approaches 100%.

When heat is taken from hydrogen during the liquefaction process, there is a slow spontaneous transition to a predominantly para ratio. This transition is exothermic, and the heat released can be a problem for the storage system. Indeed, In the case of storage at normal conditions, its conversion generates a heat of 527 kJ/kg, which is higher than the latent heat of vaporization of liquid hydrogen, 446 kJ/kg. Therefore, boil-off occurs during the storage period, lowering the overall efficiency of the storage itself.

2.3 Liquefaction plant

Nowadays, the global liquefaction capacity is about 350 tons/day. The large-scale consumers of liquefied hydrogen are aerospace industry, chemical industries, electronic/semiconductor industry and metallurgical industries. Indeed, most of the LH2 production are in the USA, because of NASA and in Canada, they installed capacity account respectively for 215 tons/day and 81 tons/day. The third LH2 productor is Japan, that follow USA and Canada with a production of 30 tons/day. Europe has currently a production capacity of 20 tons/day [8]. Nevertheless, the production capacity is expected to increase: the bigger players, as Air Liquide, Air Products, Praxair and Linde have announced the future commissioning of new plants due to the increasing demand [8].

The liquefaction process is done by cooling down the hydrogen in a gas form, usually at ambient temperature, to its liquid state at -253°C. The processes use a combination of compressors, heat exchangers, expansion engines, and throttle valves to achieve the desired cooling. The simplest example of a liquefaction process is the Linde cycle or Joule-Thompson expansion cycle, and its technical scheme is shown in Figure 4.

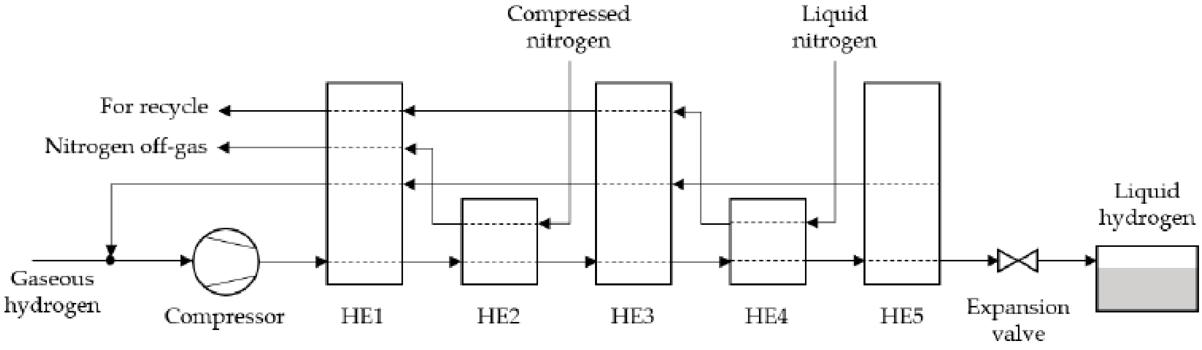


Figure 4 - Linde or Joule-Thompson expansion cycle to liquify hydrogen - image from [8]

The hydrogen before going through all the heat exchangers needs to be compressed to lower the energy needed for the cooling process. The compressed hydrogen then passes through several heat exchangers that lower the temperature step by step. The gas used to cool down the Hydrogen in the heat exchangers is usually nitrogen, since it can be cooled at room temperature thanks to the Joule–Thomson effect. The goal is then to make hydrogen reaches its inversion temperature, -73°C (the temperature at which the Joule-Thomson coefficient becomes positive). Once this temperature is reached, the -cooled gas expands in a valve and liquefies. All the gaseous form not converted is recirculated back to start again the cooling process [8].

Conversion of ortho- to para-hydrogen is very slow and can occur when the liquefaction process is already done, limiting the efficiency of the liquefaction plant itself and the storage. To prevent this loss, a catalyst is used during liquefaction to accelerate the change of ortho- to para-hydrogen; in this way the heat released is lower and the overall efficiency higher.

The liquefaction plant is a key component of the liquid hydrogen supply chain, and it has both large operational and capital expenditures. This is why it is worth going more in detail about its features, capital costs and future trends.

The theoretical work needed by the liquefaction strongly depends on the initial pressure of the energy carrier, but generally is confined in a range between 2,8 and 3,3 kWh/kg. To get an idea of this value, the ideal work of liquefaction for nitrogen is only 0.207 kWh/kg, one order of magnitude less [9].

Considering nowadays normal capacities size, between 10 and 15 tons/day, the energy required by the liquefaction, according to several references ([5], [6], [8], [10], [11]) is indicated as something in between 10 to 13 kWh/kg. Nevertheless, according to the same references and some others ([12], [13]), the electricity required is going to be lower and lower, increasing the size of the plants. For plants with a capacity of several hundreds of tons per day, the lower values expected are in the range of 5.30–8.50 kWh/kg. According to Integrated Design for Efficient Advanced Liquefaction of Hydrogen (IDEALHY), there are 3 efficiency-drivers when a liquefaction plant is concerned:

- Increasing plant scale
- More efficient process design
- Using more efficient components

According to the size of the plant, IDEALHY suggest a hypothetical trend for the energy required, that is shown in Figure 5.

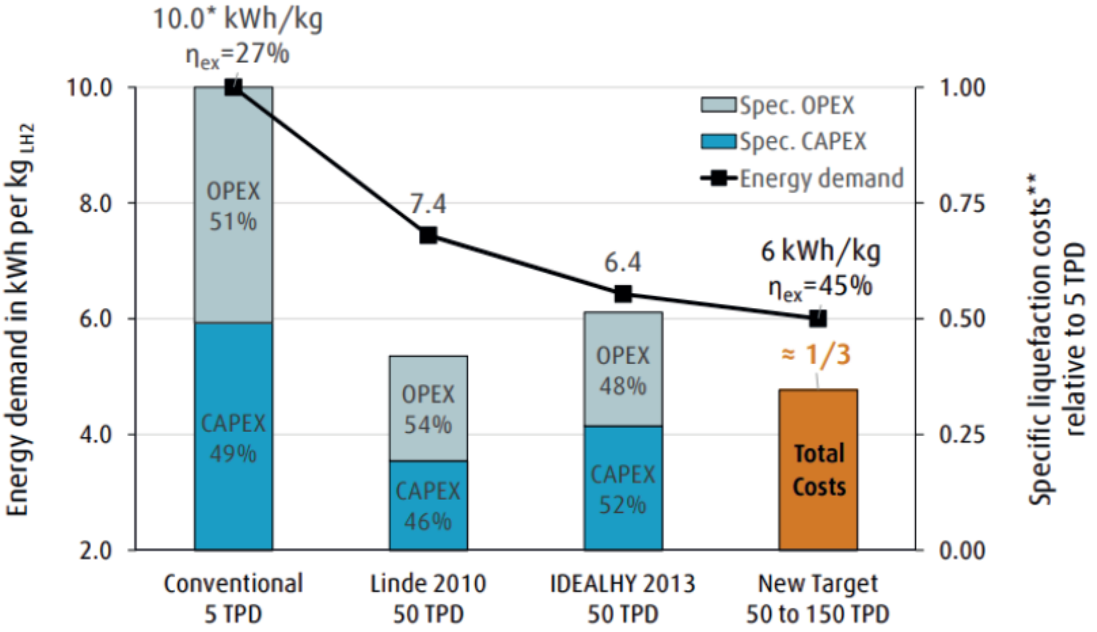


Figure 5 - Summary of Current and Projected Liquefaction Costs & Efficiencies – image from [5]

It is interesting to compare what IDEALHY indicates for the energy consumption, with the results that come from the interpolation of the literature data ([5], [6], [8], [10], [11], [12], [13]) (Figure 6).

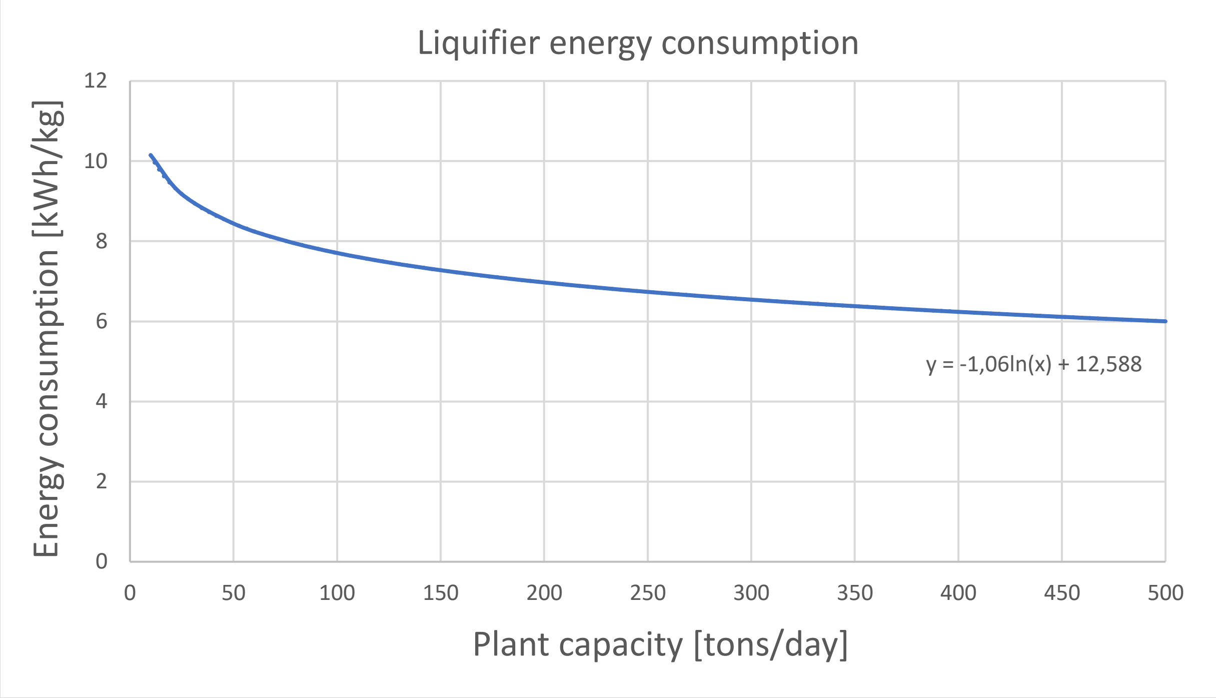


Figure 6 - Liquefier energy consumption trend according to the plant size

As IDEALHY predicts, the data from other sources seems to go in that direction too.

The capital expenditures of the liquefaction plant are among the main cost drivers in the supply chain. Depending on the size of the plant, the cost can start from few millions up to hundreds of millions´ euros. Unfortunately, no big plants are currently built; therefore, all the data found for capacities higher than 20-25 tons/day are forecasts and not real values. However, they showed a high consistency. In Figure 7 - Liquefaction plant investment cost depending on the size of the plant, the resulting fitting line and its equation from the interpolation of the literature data ([9], [10], [12], [13], [14], [15], [16]) is shown.

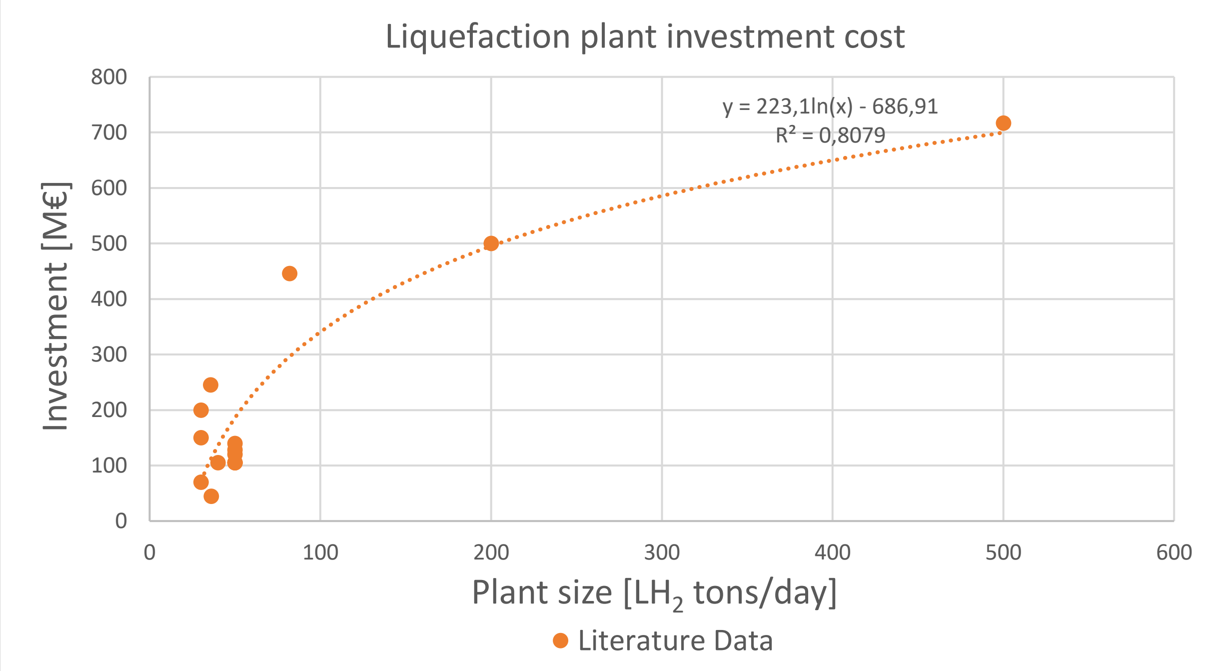


Figure 7 - Liquefaction plant investment cost depending on the size of the plant

Being just projections, the future values for bigger plants can be subject to important oscillations. For example, a plant with a capacity of 500 tons/day has a production rate of liquid hydrogen larger than the global daily capacity. In the next few years big plants are going to be built, and then more precise data will be available.

For medium-size (20 -80 tons/day) plants, instead, the literature offers more values to rely on.

2.4 Liquid Hydrogen Storage

Almost all of the knowledge in the field of storing large quantities, comes from the Aerospace industry. The largest single cryogenic storage tank in the world belongs to NASA in Florida, USA (Figure 8). This tank has a capacity of 3800 m_3, equivalent to 270 tons of LH2. Nasa is going to construct an even bigger storage with a with a capacity of 375 tons liquid hydrogen. According to [5], the future size of LH2 storage tanks will be about 13 times bigger than the NASA one, reaching capacities up to 3 500 tons hydrogen.

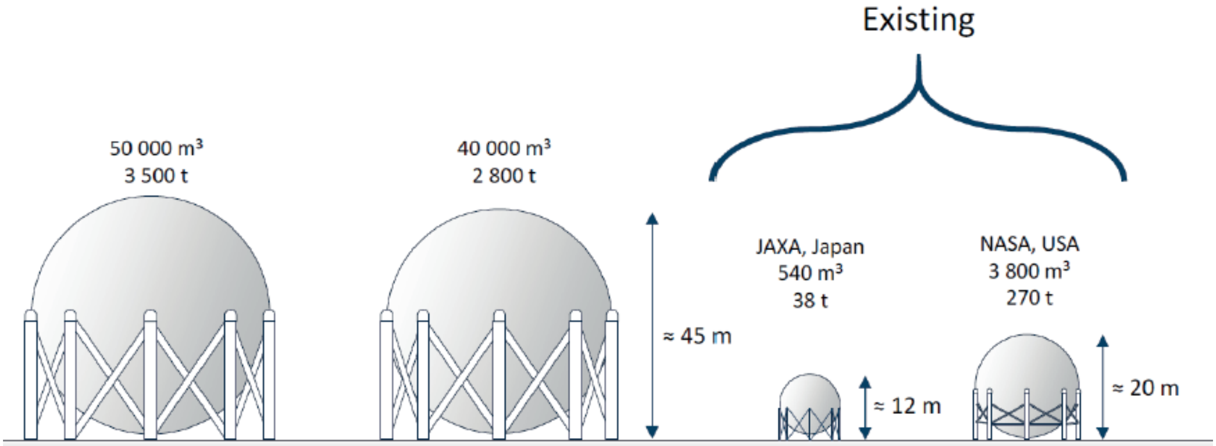


Figure 8 - Existing liquid hydrogen storage design - image from [5]

The most used shape for the storage is the spherical one. This configuration has the lowest surface area for heat transfer per unit volume. This means that as the diameter of the tank increases, the volume increases faster than the surface area; larger tanks will have then proportionally less heat transfer area than smaller tanks. In Figure 9 is shown the typical scheme of the insultation of a liquid hydrogen cylindrical storage: within the outer and the inner tank (double wall) it has been adopted a perlite vacuum insultation system; the perlite, that are an insultation material, enhance the thermal insulation in addition to the vacuum [17].When the time of storage is longer, additional outer wall with the space filled with liquid nitrogen is added. This reduces the heat transfer by lowering the temperature difference, which drives the heat transfer [8]. Anyway, even with careful insulation, some hydrogen will evaporate. Therefore, Boil of Gas (BOG) losses are unavoidable and can be just lowered. Current values for the BOG are in the range of 0,2 and 0,3% per day.

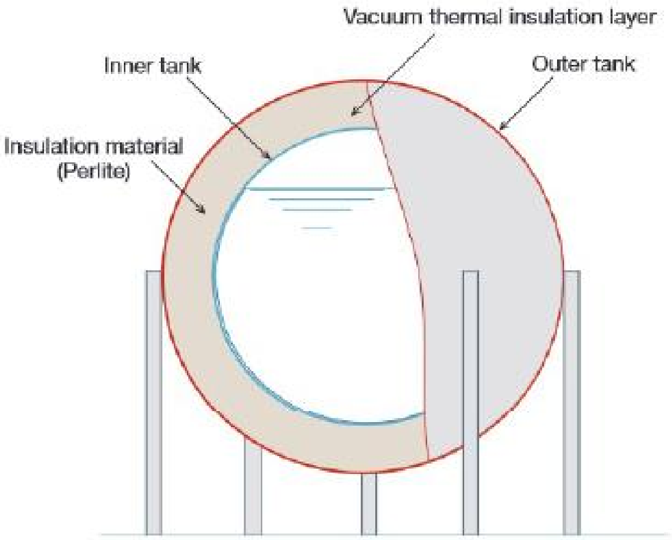


Figure 9 - Liquid hydrogen spherical storage scheme - image from [17]

Operational expenditures of LH_2 storage are related to the type of tank used. Depending on the time of storing, the storage can be kept cold using a refrigeration system or not, increasing or decreasing the overall need of energy. Anyway, since the storage in this analysis will be part of the import and export terminal its energy consumption will be included in their OPEX.

The capital expenditures of the LH2 storage are referred to the cost of the materials and equipment needed and have an important role on the final levelized cost of hydrogen; therefore, it is worth to look more in detail at their value.

For low capacities (until 10000 m^3) the literature is mainly based on the aerospace industry data, while new scientific papers and research are appearing more often year by year because of the rising commercial demand of LH2.

In Figure 10 is showed a comparison between the fitting line that comes from the interpolation of the literature data found ([5], [9], [18], [19], [2], [20]) and a curve obtained using an economy of scale approach, with a scale-up factor “n” equal to 0,85.

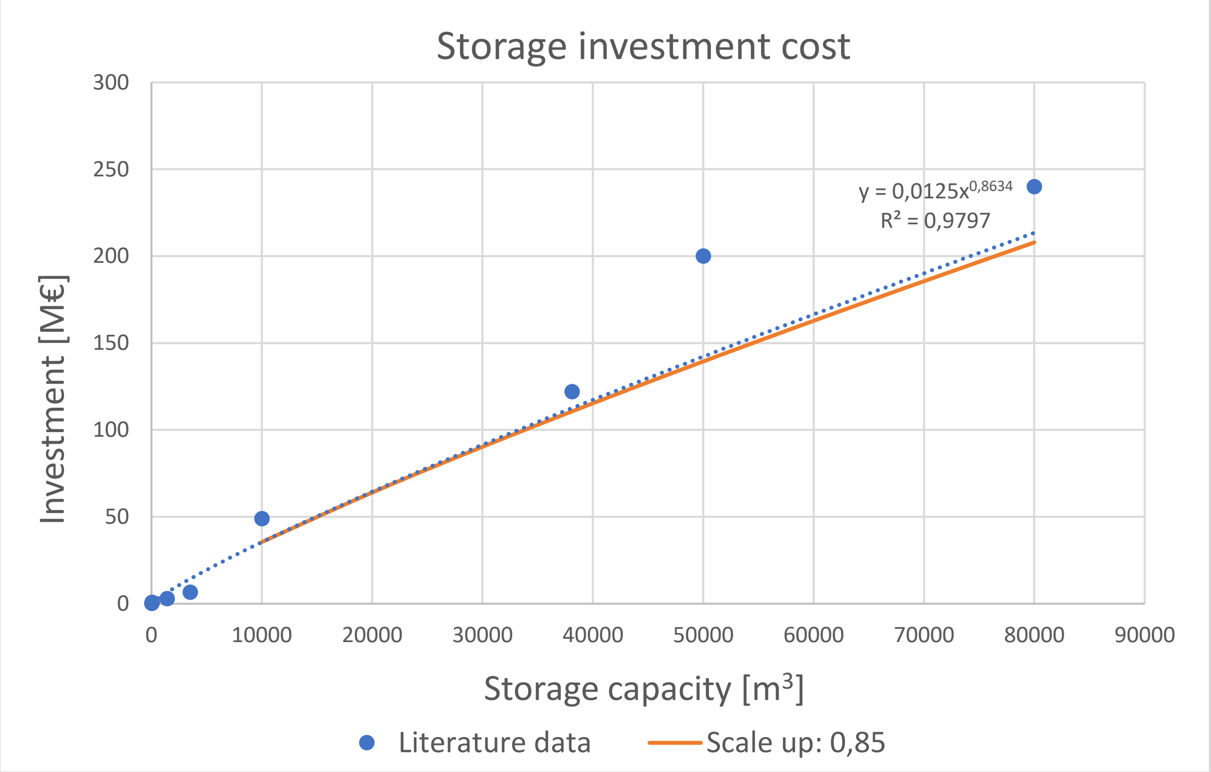


Figure 10 - Storage investment cost depending on the capacity

The economy of scale approach is used when some data regarding the size and the price of the system are available only referring a smaller size respect to the one being investigated. Given these data (S_0 and C_0), is possible to get the desired value C_1 by means of equation (2.1), once the scale-up factor is assigned. This factor “n” depends on the category the system fits in: for industrial or chemical plant is common to use a scale-up factor equal to 0,67.

$$C(S_1) = C_0 \left(\frac{S_1}{S_0}\right)^n \tag{2.1}$$

The lower the coefficient, the higher the economic gain that comes with large sizes of plant.

Therefore, the one shown in Figure 10 is an interesting behavior since, for this kind of system, the literature usually suggests using a factor equal to 0,67, but the system behavior is way far from this approximation. This can be explained by the low maturity of such technology. Anyway, sizes larger than 80000 m^3 for the storage will be unlikely, because the system would be too big, for this reason smaller but more storages will be built in that case to reach the required storage capacity.

What has been discussed until now will be useful and matter of discussion also for the next chapters.

2.5 Liquid Hydrogen tanker

The LH_2 tanker ship is probably the most unclear issue over the all LH_2 supply chain. Worldwide, there is only one small prototype designed by Kawasaki: long 116 meter and with an overall capacity of 2500 m^3 or 180 tons of hydrogen [5]. Therefore, in the literature is almost impossible find data about it, either because they are still industry secrets or because there are not yet. Kawasaki is the company that is investing more on LH_2 tanker, and after the release of the prototype (Figure 11) is scheduled the realization of a ship that will carry 160000 m^3 of liquid hydrogen.

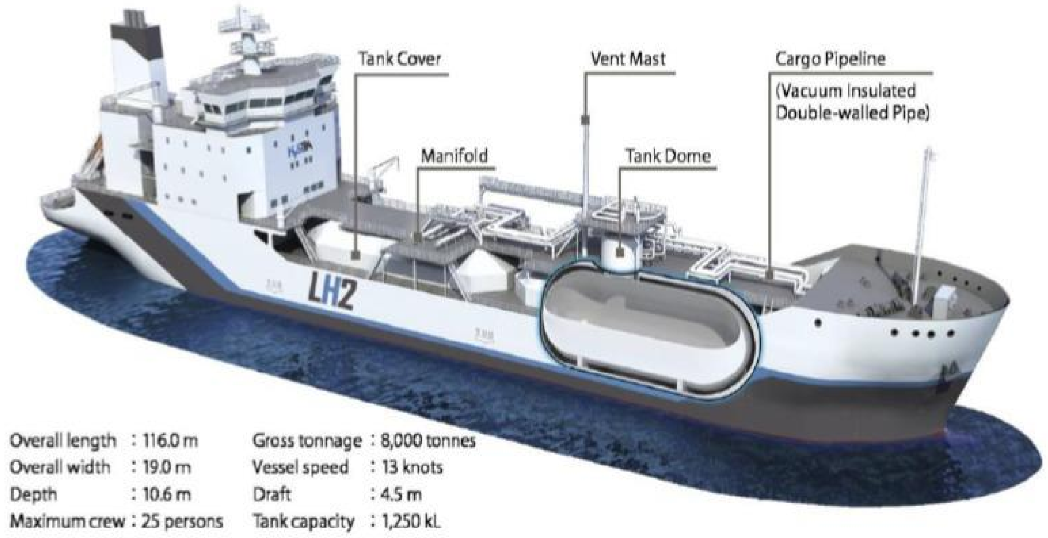


Figure 11- Outline of a Pilot LH2 Carrier Ship - image from [21]

Nevertheless, the technical challenges in designing a LH_2 tanker are not few. According to [22], the low density of liquid hydrogen (71 kg/m^3) compared to the density of LNG (450 kg/m^3), results in a ship of very large dimension with respect to the weight of the cargo it carries. Such a light weight for a ship long more than 300 meters implies an unusually shallow draft for a ship of those sizes: the DWT is slightly larger than 20,000 tons, and the difference in the draft between the loaded and unloaded condition is less than 1 meter.

Given these operative conditions, [22] suggests what would be the best configuration for a LH_2 tanker ship. [22] indicates four tanks as best design choice instead of a unique one, (Figure 12), because it allows to maintain a better equilibrium of the hull of the ship while sailing.

In both the prototype and the final model designed by Kawasaki, the engine is a diesel motor, because hydrogen based marine power system are still not mature. However, Kawasaki is studying the possibility of exploiting the boil of gas losses to supply the engine, as it currently happening for the LNG ships. In doing so, also the thickness of the LH_2 tanks on board will be sized accordingly to the need of the engine. The power generation, based on typical engines used for LNG´s ships of the same dimension, is assumed to be 50 MW to ensure a service speed of 18 knots [22].

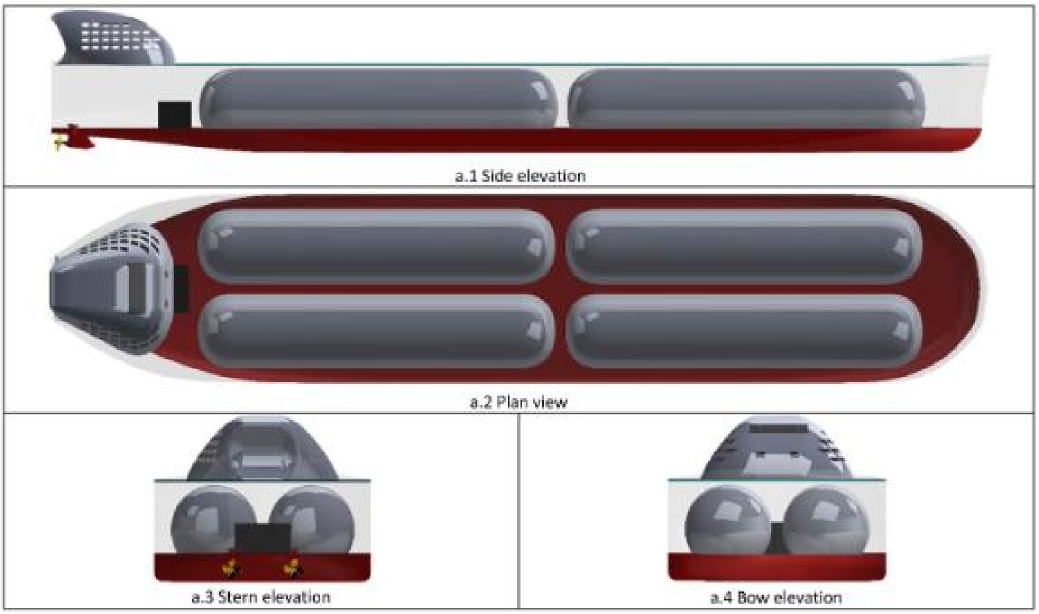


Figure 12 - Hypothetical LH2 carrier scheme for big quantity transported - image from [22]

2.5.1 Tanker's cost evaluation using the literature approach

The interpolation of the data found on the literature ([12], [14], [23], [24], [25]) led to the cost function of the ship investment shown in Figure 13. On the same graph, is plotted a curve with a hypothetical scale-up coefficient of 0,5 (red line) as it has been done already for the storage.

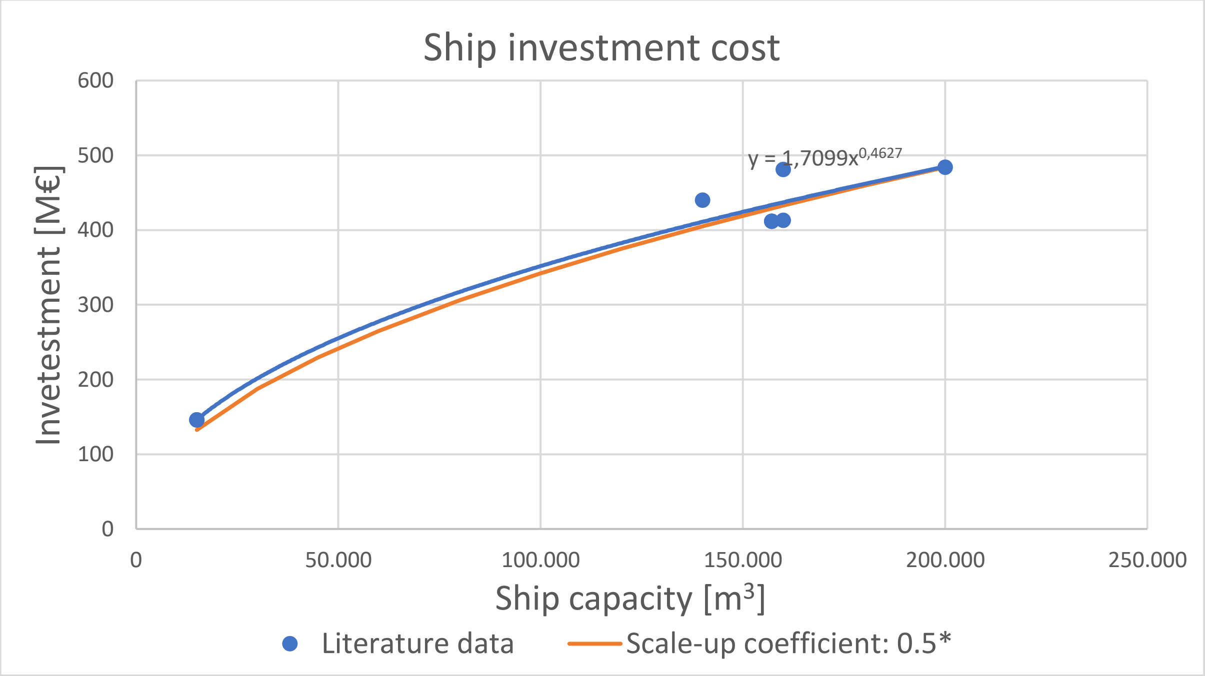


Figure 13 - Ship investment cost according to the ship capacity

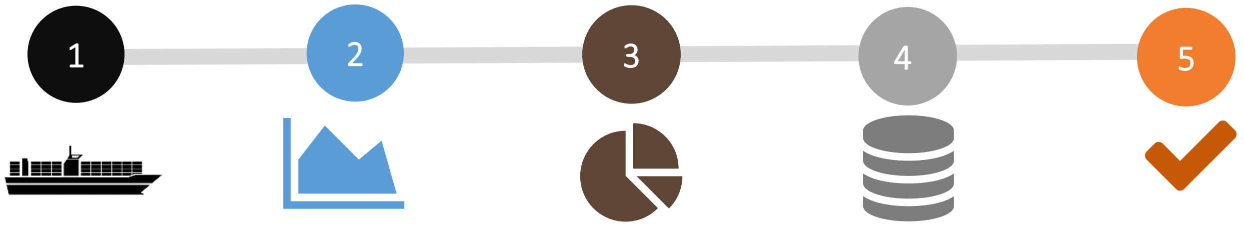
The literature cost function and the curve with scalability behavior tend to overlap, and this is surprising because the scale-up coefficient used is low (0,5). Indeed, there is a mismatch between what has been seen for the storage cost trend and the ship cost trend. For storage, which is one of the main ship's cost drivers, the scale-up coefficient was equal to 0,85, meaning that there is not such a great economic gain when scaling up a system, as instead Figure 13 suggests for the ship.

The storage used on board is the same of the one used in land, plus some additional safety requirements due to the not-static condition related to the navigation, which surely will not make the price lower; this difference is not justified anywhere. Because of this mismatch on the scalability of the system, another approach is proposed in section 2.5.2. In this case, instead of considering the ship as a whole system as the literature did, the cost of the ship will be evaluated, considering it as the sum of two separated units: the bare ship, without storage, and the storage.

2.5.2 Tanker's cost evaluation adopting a new approach

For the new approach, the starting point is assuming that the bare LH_2 tanker is comparable with the LNG tanker, as most of the literature suggests. To get then the final cost, it will be necessary to add to the bare ship the cost of the liquid hydrogen storage that follow the cost function showed in section 2.4.

The whole process is shown below, and it consists of 5 steps. The first step (1) is finding enough values to get a cost function that relates the LNG tanker capacity and its cost; this cost function is shown in Appendix I. Once the desired capacity is known the price of the LNG tanker ship can be obtained from (2). At this point, from this value the cost of the LNG storage (3), which usually is in the range of 40-45% of the ships´ cost, it has to be subtracted (4). Finally, the last step consists in adding the cost of the storage with the desired capacity, to the cost of the bare ship (5).



1 LNG tanker literature review

12 LNG ship´s cost found on scientific papers for different LNG capacities [Appendix III]

2 LNG tanker cost function

Fitting curve based on the literature review:

$y = 0,4459x^{0,5094,}$

$R^2 = 0,9698$

3 **LNG storage cost evaluation**

LNG storage cost assessment

4 Bare ship cost = Tanker Investment (2) – Storage cost (3)

Evaluation of the LNG ship´s cost without storage

5 **LH2 ship cost = Bare ship cost (4) + LH2 storage cost**

Low BOG ship´s cost found added the on-shore storage price. Same storage used at the terminal

In Table 3, 3 examples with different storage capacities are reported using this approach. The final values found with this approach will be compared with the results coming from the literature approach in the next section.

Table 3 - LH2 ship cost evaluation examples- bottom-up approach

Capacity [m³]	LNG ship´s cost [M€]	LNG storage cost [M€]	% storage / ship	Bare ship cost [M€]	LH2 storage cost [M€]	LH2 ship cost [M€]
28400	82,7	37	44,7	45,7	110	155,7
114000	168	69,25	41,2	98,75	380	478
166000	203,7	84	41,3	119,7	558	677,7

Having a lower scale-up coefficient the literature approach guarantees, compared to the new approach, a lower price when it comes to larger capacities. In Figure 14 are shown both the approaches depending on the number of ships adopted. The annual hydrogen production is fixed to 500 tons/day in Figure 14, therefore more ships means lower capacities; as can be seen in Figure 14,, for the 1 ship case the literature approach (red line) indicates a value (493 M€) that is way lower than the new approach (Blue line) value (719 M€) and this cost´s difference gets lower until when the bottom-up approach becomes cheaper for lower ships´ sizes. Indeed, as lower capacities are reached the influence of the scale-up coefficients becomes less important.

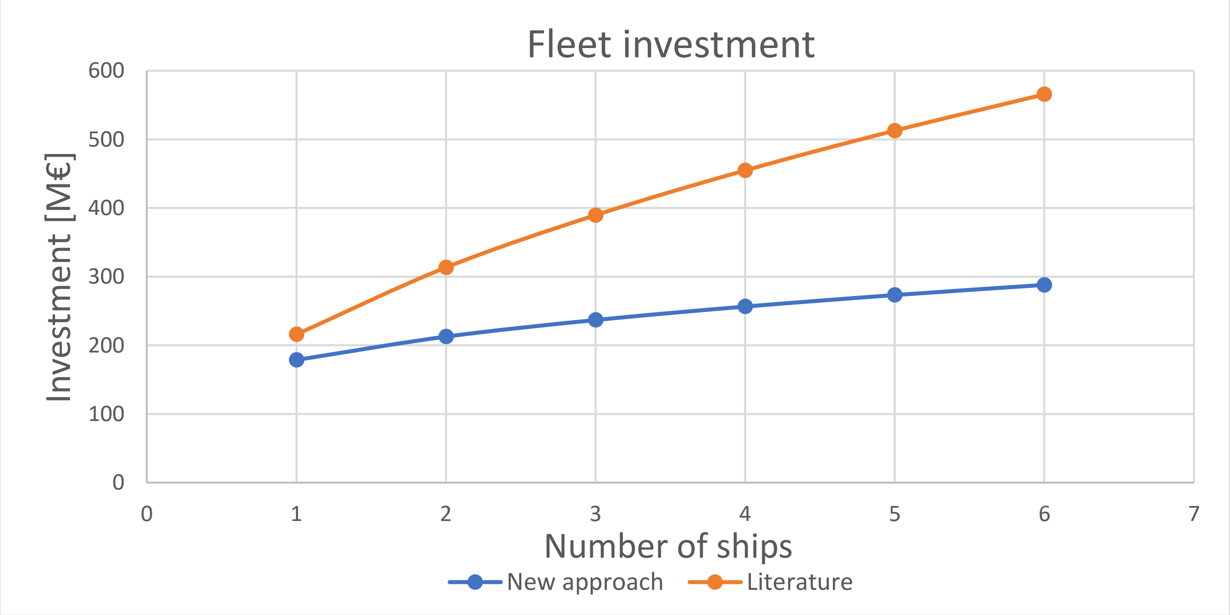


Figure 14 - Fleet investment cost. Comparison between approaches

In Table 4 are listed the main hypotheses made for the evaluation of the LH_2 ship´s O&M.

Table 4 - Liquid Hydrogen ship´s hypotheses

Hypothesis	Value	Unit
Boil Of Gas during loading	5,60	%/loading
Boil Of Gas during sailing	0,3	%/day
Boil Of Gas during unloading	6,10	%/unloading
Fuel Price	20	€/GJ
Fuel Consumption	2500	MJ/km
O&M	4700	€/day

2.6 Import and Export terminals

The liquefied hydrogen terminal consists of a liquefied hydrogen storage tank for storing liquefied hydrogen, crane, and pumps to load and unload liquid hydrogen between a carrier and the shore, ancillary facilities like piping, valves and fitting, plus specific facilities like the BOG handling systems, offices and buildings. Whether an import or an export terminal, the design does not change exception made for the terminal's capacity: import terminals are usually bigger because they are linked to several trade routes, accommodating several carriers.

However, finding values about the terminals´ cost is not an easy task. [7] and [23] respectively give a value of 878 M€ and 290 M€ for terminals with a capacity of 200 000 m^3 and 46 000 m^3 each; but these 2 data are not enough to get a function that give back the cost of a terminal given the capacity of liquid hydrogen that can be handled in it. [7] suggests using the already discussed scale-up coefficient 0,67. Nevertheless, as already explained in section 2.5 for the LH_2 tanker, adopting this approach would be a bit too optimistic at this stage of the technology. Next section 2.6.1 will be dedicated to a bottom-up analysis, trying to characterize the system features, and assigning them cost functions.

2.6.1 Bottom-up analysis of the system

In the bottom-up analysis the main features of the terminal will be investigated in the attempt to assign proper cost functions that will help in a second moment to the evaluation of the capital expenditure for the whole system. In Figure 15 is possible to see which can be the relative weights of each component; the example of Figure 15 is made for the specific case of 500 tons/day hydrogen production, 3 ships fleet and 10000 km of transport distance.

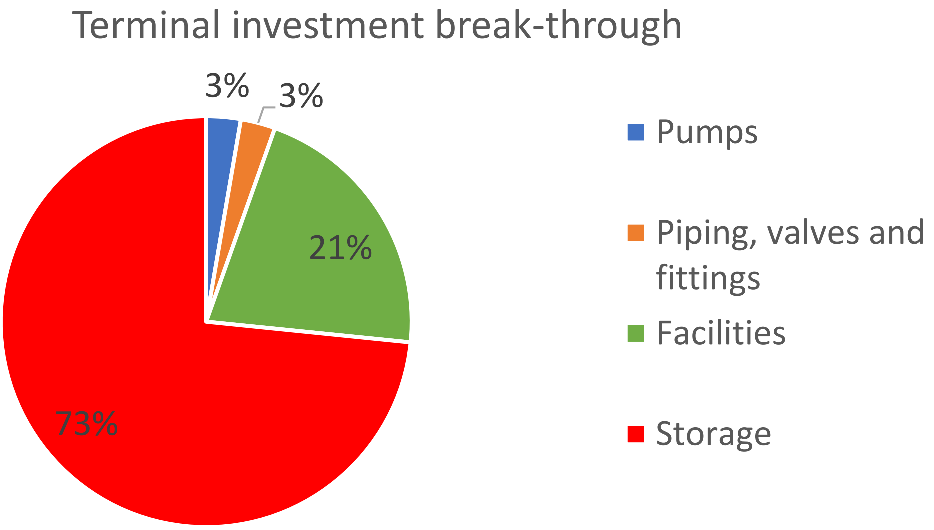


Figure 15 - Terminal investment break-through

The storage has the largest contribution on the overall investment cost (65-75%). The storage type for the terminal is of the same type already analyzed in section 2.4. Having this great influence on the terminal´s costs, the approach proposed by [7], is once again too optimistic.

The facilities can include the terminal offices, operative buildings and the system that handles the BOG flow: both the terminal of import and export having storage tanks face unavoidable losses. The hydrogen that comes back to the gaseous form, can be liquified again, stored as it is or compressed and stored, therefore a unit that handle the loss is necessary. Since the facilities cost is specific of each system, the cost function in this case will follow the scaling approach suggested by [7].

The pumps compared to other systems have not a secondary effect on the final price, because of the extreme low temperatures they have to bear, and the low heat transfer allowed from the environment to the liquid. Cryogenic pumps are currently not ready to face big flow quantities (over 50 tons/h) required by such big terminals [10]. Therefore, it will be assumed that the pumps will be installed in parallel, summing the cost of the singular pump. The cost for a pump is given by [26].

The piping, valves and fitting investment depends on the dimensions and capacity of the terminal, and their cost function is given by [26] and can be seen in Appendix I.

Once the cost functions of the terminal´s components are assigned, the sum of their values will give back the overall terminal investment cost. In Figure 16 is showed the comparison between the cost functions that come from the literature scalability approach (orange line) and the bottom-up approach (blue line).

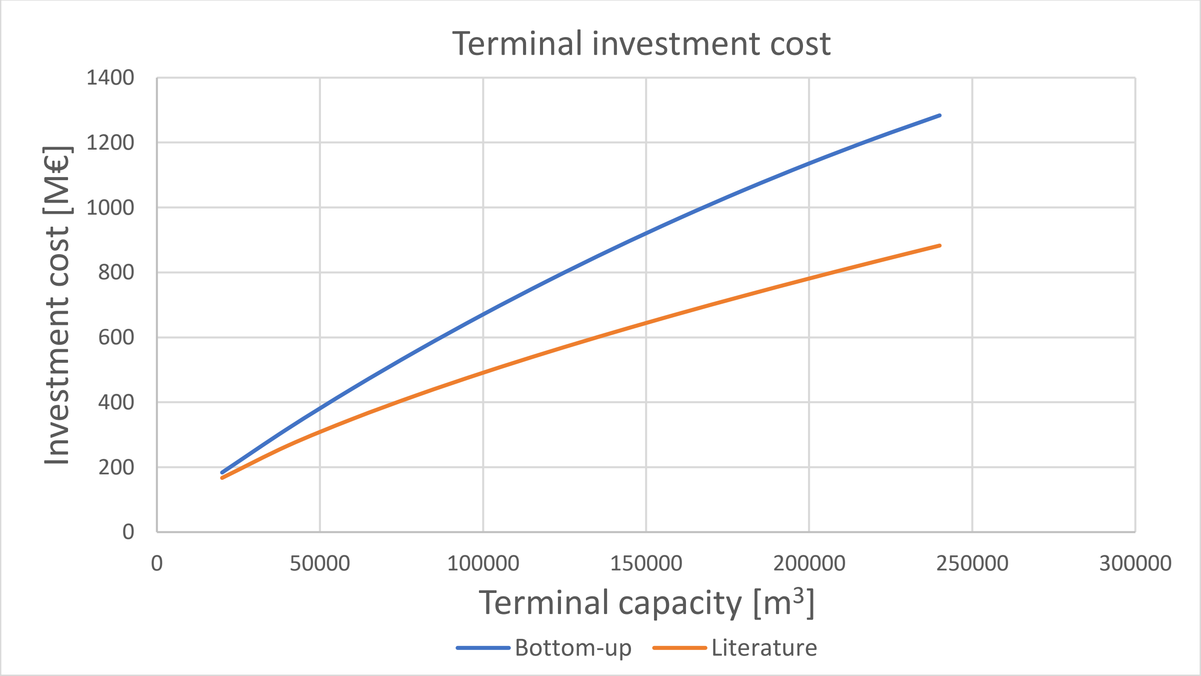


Figure 16 - Terminal investment cost. A comparison between the bottom-up and literature approach

As it was easy to expect, adopting an economy of scale approach will lead to a save on the investment cost especially for high capacities. In any case, this analysis does not want to indicate which method is more valid, but rather wants to propose a different point of view, less optimistic. The final price could lie in between these two approaches. In Table 5 are instead listed some of the hypotheses made for the evaluation of the terminal OPEX.

Table 5 - Liquid Hydrogen terminals´ hypotheses

Hypothesis	**Value**	**Unit**
Terminal oversizing factor	1,3	-
Electricity Consumption	198	kWh/tH_2
Electricity Price	18	€/MWh
Annual O&M	2	%/total CAPEX

2.7 Levelized Cost Of Hydrogen (LCOH)

The Levelized Cost Of Hydrogen (LCOH), is evaluated dividing the sum of the annualized CAPEX and yearly OPEX by the annual ammonia production traded. The yearly LH_2 delivered is equal to the product of the daily liquid Hydrogen production for the operating days of the system minus the BOG losses of the supply chain (loading, shipping, unloading and storage). The LH_2 in gaseous form at the import and export terminal, as a conservative assumption, is considered a loss, not involving the possible positive income coming from the selling of it. The mathematic process to get the LCOH is shown in equation (3. 1) and (3. 2).

$$LCOA = \frac{\sum_i \frac{CAPEX_i}{a.f._i} + OPEX_i + \ O\&M_i}{LH2\ treated\ annually}$$

(3. 1)

Where the actualization factor $a.f.$ is equal to:

$$a.f. = \frac{\left(1 - \left(\frac{1}{\left(1+\frac{r}{100}\right)^{l.t.}}\right)\right)}{\left(\frac{r}{100}\right)}$$

(3. 2)

With:

- r = discount rate
- l.t. = lifetime [years]
- LH_2 treated annually = [kg/year]

Based on equation (3. 1) and (3. 2) the LCOH has been evaluated and then displayed in a pie chart that shows the relative weights of the LH_2 system´s main features. Figure 18 shows the percentage for the new approach, while Figure 17 shows the percentage according to the Literature approach.

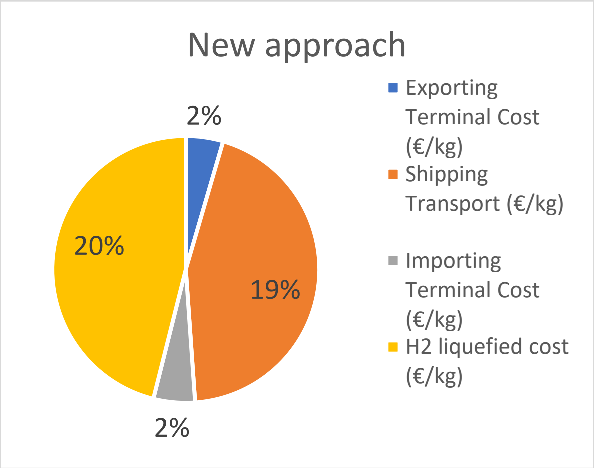


Figure 18 - LCOH break-through, new approach

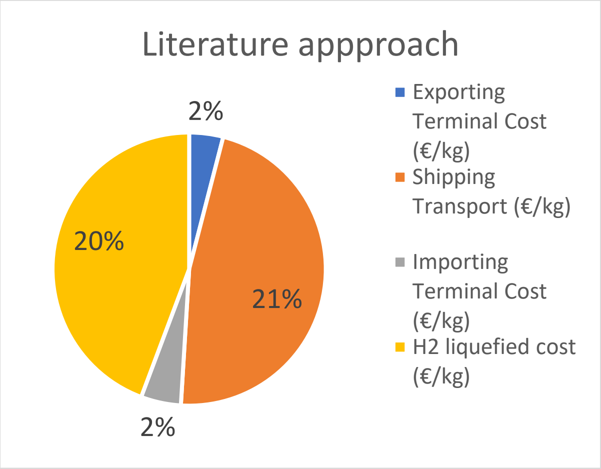


Figure 17 - LCOH break-through, Literature approach

The two LCOH breakthrough in Figure 18 and Figure 17 are referring to a specific case, where the daily production is equal to 500 tons per day, the transport distance is equal to 10 000 km and the fleet consists in three ships. In this case the difference between these two approaches is not so relevant, with a LCOH equal to 2,04 for the new approach and 1,94 for the literature. The difference will become larger for bigger ships and terminals, where the scalability coefficient will take a role.

2.7.1 Sensitivity analysis

The most important parameters that influence the LCOH are the distance of transport, the daily hydrogen production and the number of ships. Therefore, a sensitivity analysis of these parameters is done in the next pages.

The distance of transport has influence on the operative costs, like fuel and maintenance, but also on the capacity of the ship and terminal: when the daily production is fixed, increasing the distance means increasing the day of storage at the terminals which are daily refueled by the liquefaction plant. So, the more days the ship is away, the bigger they have to be the tanks. Moreover, distance becomes more important when talking about LH_2 because longer distances also imply higher BOG losses during the shipping phase.

The trend for both the literature (blue line) and the new approach (green line) assuming the case of 1 ship, is displayed in Figure 19.

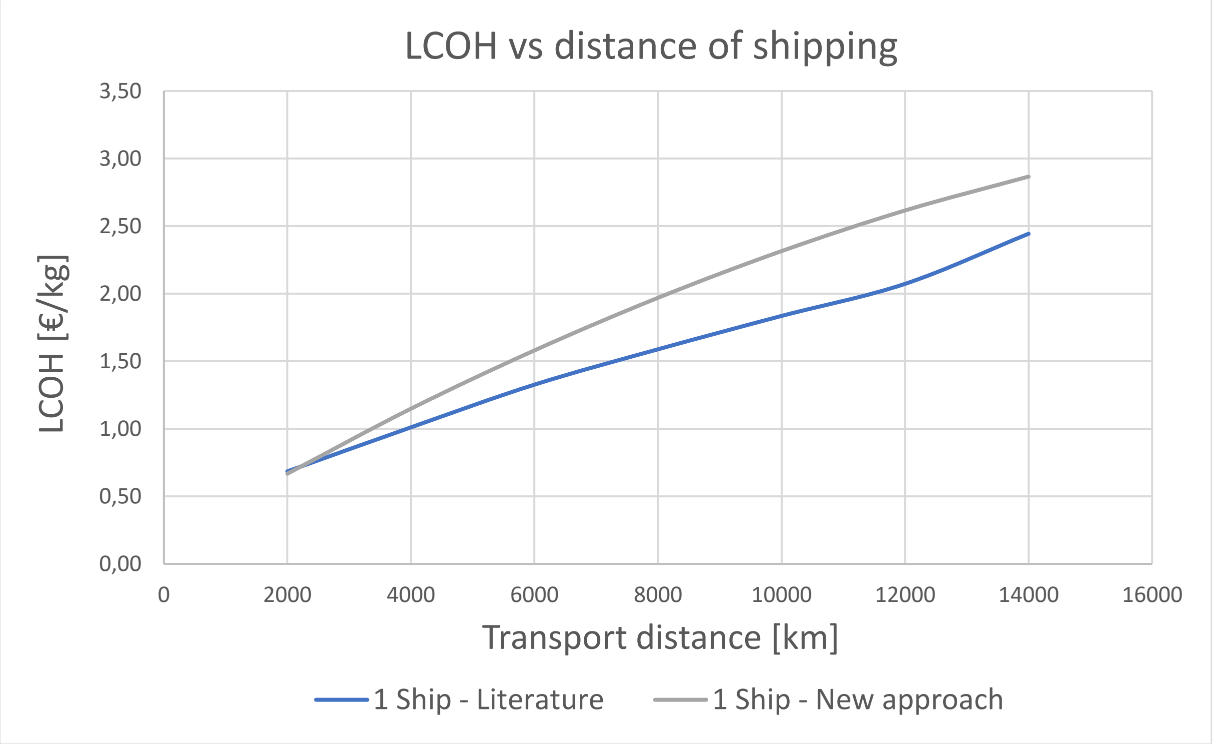


Figure 19 - LCOH for LH2 vs distance of shipping

The example of Figure 19 is made with the assumption of 500 tons of liquid hydrogen per day. The LCOH is very sensitive to the distance: for a distance of 2000 km the LCOH for both the approach is close to 0,7 €/kg. For the higher distance taken in consideration, 14000 km, the LCOH goes up to 2,20 €/kg for the literature case, and 2,87 €/kg for the new approach.

The Liquid hydrogen production, as it is shown in equation (3. 1), has a direct influence on the LCOH. Being at the denominator, when it increases, it has positive effects on the LCOH. Of course, a higher production rate needs bigger tanks and terminals, but the specific cost is lowered overall.

The choice of the number of ships is one of the key points that affect the LCOH. Indeed, the number of ships influences the size of the whole system. The higher the number of ships, the higher the fleet investment cost and fuel consumption, but on the other hand the lower the terminal capacity for both the export and import. The trade-off between these values will give the best configuration of the system.

In Figure 20 are shown the LCOH curves depending on the number of ships, fixed the daily production to 500 tons and the transport distance to 10 000 km.

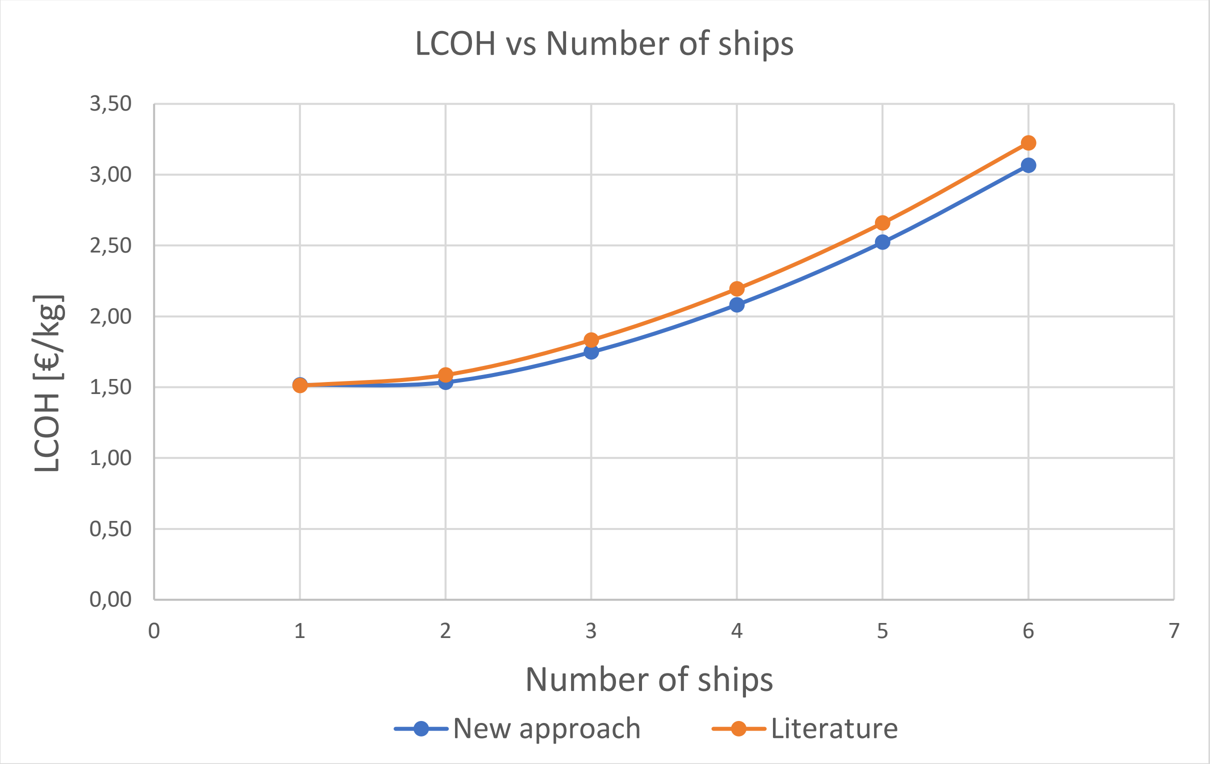


Figure 20 - LCOH of LH2 vs Number of ships

For this specific condition, the best configuration for the literature approach (red line) matches with the choice of 2 ships, leading to a LCOH of 1,48 €/kg. On the other hand, for the new approach (Blue line) the optimal point slightly changes between the choice of 2 or 3 ships, whit the best solution at 3 ships that give a LCOH equal to 1,87.
Nevertheless, the best configuration for the number of ships is not unique but concern the specific conditions of the system. Different distance of transport or daily hydrogen production could have a different optimal number of ships.

As is shown in Figure 21, while for 200 tons per day of hydrogen production (orange line), the best choice would be the purchase of just one ship, for 600 tons per day (Blue line) would be instead 2 ships. Looking at the 3 curves plotted in Figure 21 is possible to see that for lower daily production rate, a higher number of ships is less beneficial for the LCOH. Indeed, being the production lower, splitting the capacity in more ships lead to a higher fleet investment cost as it was shown in Figure 14.

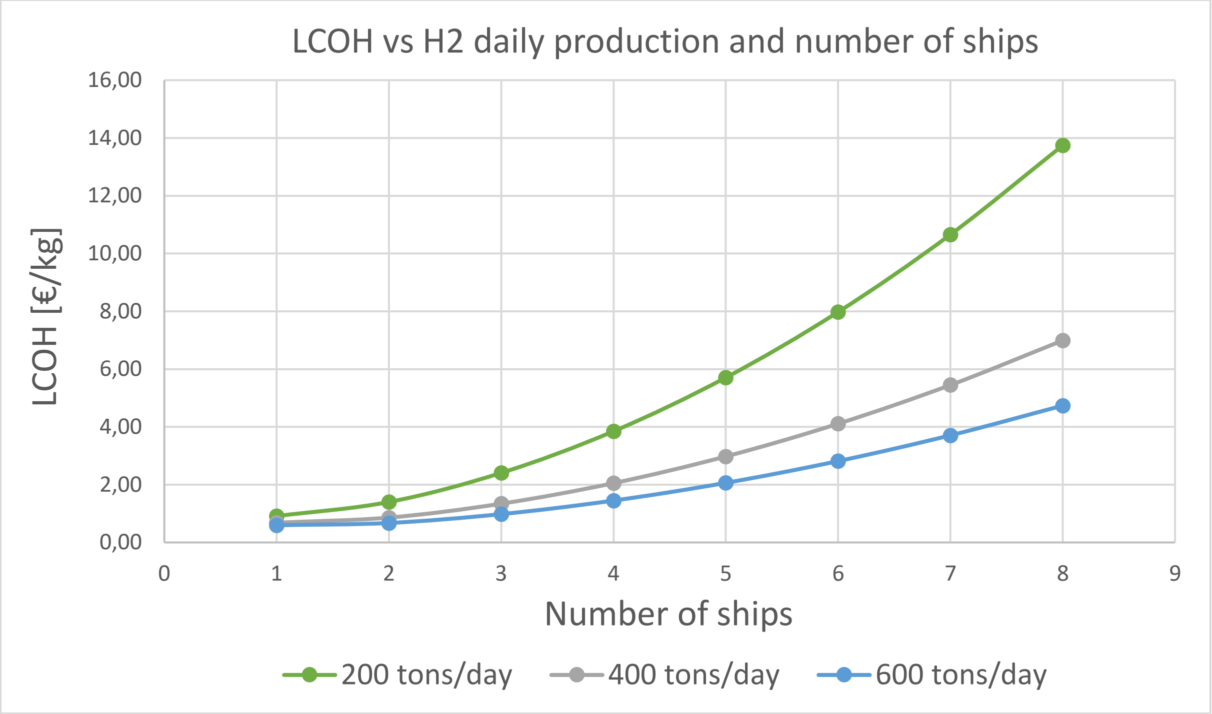


Figure 21 - LCOH of LH2 vs daily production and number of ships

Figure 21 has been obtained for the new approach for a transport distance of 10000 km. The same case, but for the literature approach is in Appendix I.

3 Transport of Ammonia as a Hydrogen energy carrier

3.1 Introduction

Ammonia, the name used for the NH3 molecule, is a compound of nitrogen and Hydrogen, which is used primarily for nitrogen-based fertilizer production, but it shows good potential for applications as a synthetic fuel in internal combustion engines and gas turbines and its possible reconversion to H2.

Nowadays, ammonia-based fertilizers feed about 3 billion people by enabling more food to be grown on a given area of land. It is estimated that the world population would be 2 to 3 billion fewer without its usage [27]. Since more than 80% of ammonia is used as fertilizer, ammonia production will increase accordingly to population growth, which is projected to rise by nearly 30% by 2050. At the same time, natural gas, that is the main feedstock for ammonia, is expected to become more expensive over the next 20 years, making the use of green Hydrogen for sustainable ammonia synthesis competitive. Furthermore, looking at the current 'geopolitics' situation and the price instability due to the war in Ukraine, decoupling the use of natural gas from fertilizers production is also something desirable. Indeed, any significant disruption in the natural gas supply can have devastating effects on the global food supply chain through decreased fertilizer production [27].

Besides in agriculture, the interest in ammonia is increasing lately due to the ease of transport and storage and because it is an energy-dense fuel with few technical barriers to adoption. Indeed, at ambient pressure it is in liquid state at a temperature of -33°C, compared to liquid Hydrogen which is instead liquid at -253°C. Therefore, it needs relatively low energy inputs to make it liquid, and also the equipment is not as complex as for Hydrogen. Moreover, the field of ammonia production and transport is already established since it is used and traded worldwide.

Unfortunately, the actual ammonia production capacity is environmentally unsustainable in the long term. Today ammonia is produced with Hydrogen from Steam Methane Reforming (SMR), a natural gas (NG) based process, and its NG consumption represent 0.5-1.0 % of the national natural gas consumption. Given this contest, the following analysis will focus on a sustainable production process of ammonia and its related transport, considering the spreading of new trade' ' routes and the need to build new infrastructures.

An overall picture of the ammonia supply chain that is going to be studied is shown in Figure 22.

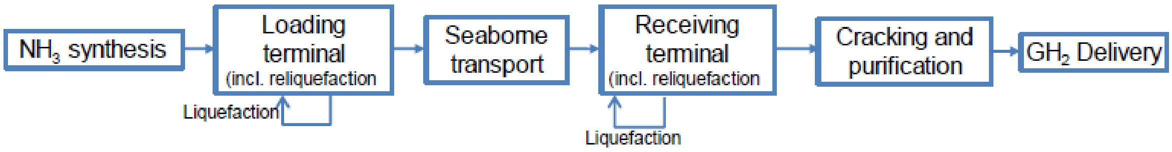


Figure 22 - Overall system of ammonia transport - image from [7]

The study will start from the sustainable synthesis of ammonia, through a classic Haber-Bosch synloop fed by green Hydrogen, then it will go through all the transport chain features, such as loading and receiving terminal and the ship itself. Finally, the possibility of cracking ammonia to get Hydrogen back will also be considered and added to the final levelized cost of ammonia (LCOA).

3.2 Physical properties

Ammonia presents similar properties to water in many reactions. Indeed, they are both diamagnetic and possess the same electron configuration. Moreover, liquid ammonia has a dielectric constant 15 times bigger than most of condensed gases, making in particularly able to dissolve many substances [28]. Some of the main properties are instead listed in Table 6.

Table 6 - Ammonia main properties

Property	Value	Unit
Liquid density (@0°C, 101.3kPa)	0.6386	g/cm3
Gas density (@0°C, 101.3kPa)	0.7714	g/L
Critical pressure	11.28	MPa
Critical temperature	132.4	°C
Boiling point	-33.43	°C
Heat of vaporization	1370	kJ/kg

Talking about its components, the only reactions of molecular nitrogen N2 at ambient temperature are the formation of lithium nitride Li3N. Above 500°C nitrogen reacts with some elements, especially with metals (nitride formation). The reaction, in absence of catalysts, between nitrogen and Hydrogen in the gas phase requires an activation energy of 230-420 kJ/mol at temperatures of 800-1200 K to achieve measurable reaction rates. In the catalytic combination of nitrogen and Hydrogen, the molecules lose their translational degrees of freedom by fixation on the catalyst surface; this leads to a sensible reduction of the activation energy required to 103 kJ/mol and a reaction temperature range of 250 – 400 °C [28]. At pressure above 200 MPa, the synthesis of ammonia proceeds even in the absence of specific catalyst.

Finally, according to Figure 23, the lower the temperature and the higher the pressure inside the synthesis loop, the better the yield of the reaction.

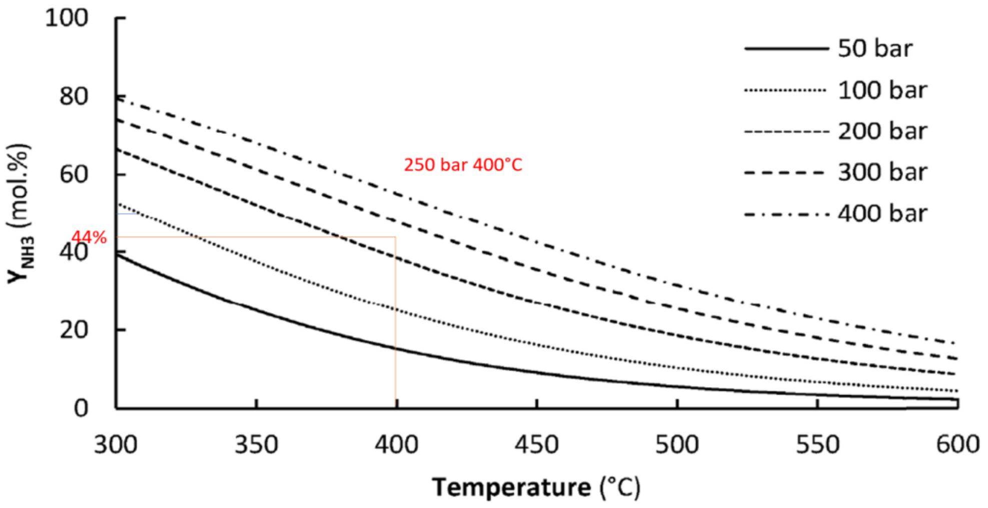


Figure 23 - Ammonia synthesis yield depending on the pressure and temperature of the synthesis – from [3]

3.3 Ammonia market

The ammonia market was estimated to be around 178,330 kilotons in 2021 [29], a value pretty stable in the last decades, without any relevant peaks. From an economic point of view, the ammonia market has been evaluated for \$71.98 billion in 2021 and it is expected to grow up to \$110.93 billion in 2028 [30].

Ammonia can be used for several applications, as can be seen in Figure 24, but for nearly 80% of the end-user industry, ammonia is used as fertilizer in the agriculture field.

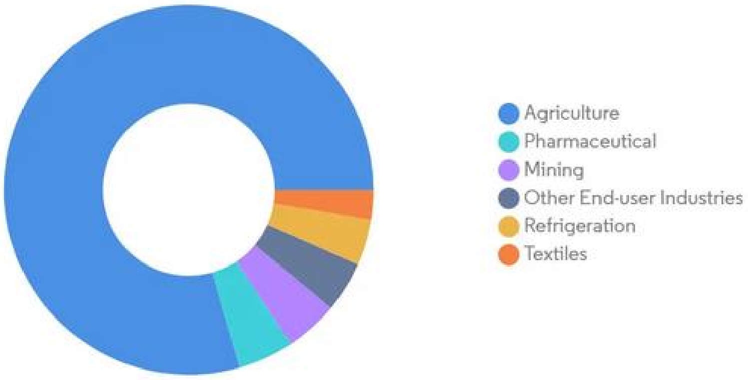


Figure 24 - Ammonia market, volume % by end-user industry. Global 2021 - image from [4]

Looking at future trends, the largest growth is projected to be in South Asia, Latin America, and Southeast Asia. This can be explained by the increasing rate in the population growth, which, as already said, will boost the ammonia demand to satisfy the food request.
But the agriculture field is not the only one expected to grow in the next decades; according to [31] the shipping sector would require five times as much ammonia as the current global fertilizer industry by 2050 if transition to 100% ammonia will happen successfully.

In addition, if the ammonia takes always more importance as a way of hydrogen transport, the demand will grow sensitively compared to the current demand.

3.4 Ammonia synthesis plant

Today ammonia is the second largest synthetic chemical product and more than the 90% of its world production is synthetized from nitrogen and Hydrogen in a catalytic process. The first to experiment such process have been Frity Haber and Carl Bosch using a promoted iron catalyst discovered by Alwin Mittasch almost one century ago, the process has been named after them, Haber-Bosch process. Since the early days there has been no fundamental change in this process: a hydrogen-nitrogen mixture reacts over an iron catalyst at elevated temperature, in the range of 400-500°C, and pressures above 100 bar [28].
The complete process of industrial ammonia production may be subdivided into the following sections:

- Synthesis gas production
- Compression
- Ammonia synthesis

The synthesis gas production concerns the Hydrogen and nitrogen production: the Hydrogen used in this reaction, nowadays mainly comes from steam reforming processes (85% of the total production), but in order to be sustainable has to be produced from electrolysis [32]; this process usually takes place far from the ammonia plant and so Hydrogen needs to be transported to the synthesis plant afterward. The other component, nitrogen, is instead produced at the plant site by means of an Air Separation Unit (ASU).

The compression stage is mandatory to match the conditions required by the catalytic synthesis, but ensuring lower temperatures has the consequence of less energy to be used into the process.

For the synthesis, the goal is preparing a pure mixture of nitrogen and Hydrogen in the stoichiometric ratio of 1:3 [28] and letting it reacts under the right conditions of temperature and pressure to ensure the right velocity of reaction, following the equation (2.1).

$$N_2 + 3H_2 \leftrightarrow NH_3 \qquad \Delta H = -92.44\, kJ/mol \tag{2.1}$$

This reaction is usually done with the help of an iron-based catalytic bed and is exothermic, meaning that heat is released from the creation of bonds for the ammonia molecule. The reaction, however, 'doesn't convert the entirety of the mix into ammonia. This is the reason why, after the separation of NH3, the exiting N2 and H2 has to be recycled in the process, resulting in what is known as the Haber Bosch synthesis loop which is sometimes also called ""synloop"".
When talking about the synthesis facility plant, the main features are the ammonia converter (usually a Haber-Bosch synloop) and an ASU (Air separation unit) which depending on the capacity and N2 purity required can be: a membrane for low capacities, a PSA system for medium capacities and a cryogenic distillation for high capacities [27].

The typical size of a large ammonia plant is in the range of 1,000-1,500 tons per day, although capacities up to 1,800 tons per day and above are not uncommon for new plants [32] and even higher capacities are expected to be built in the upcoming years.

3.4.1 Air Separation Unit (ASU)

For the source of nitrogen, an air separation unit is used, being the composition of air made of 78,08% nitrogen, 20,95% oxygen, 0,93% argon and a small amount of other gas. There are different types of separation methods with their own advantages and disadvantages. In Figure 25, it is possible to see which are the common ranges in which every technology of separation has its focus.

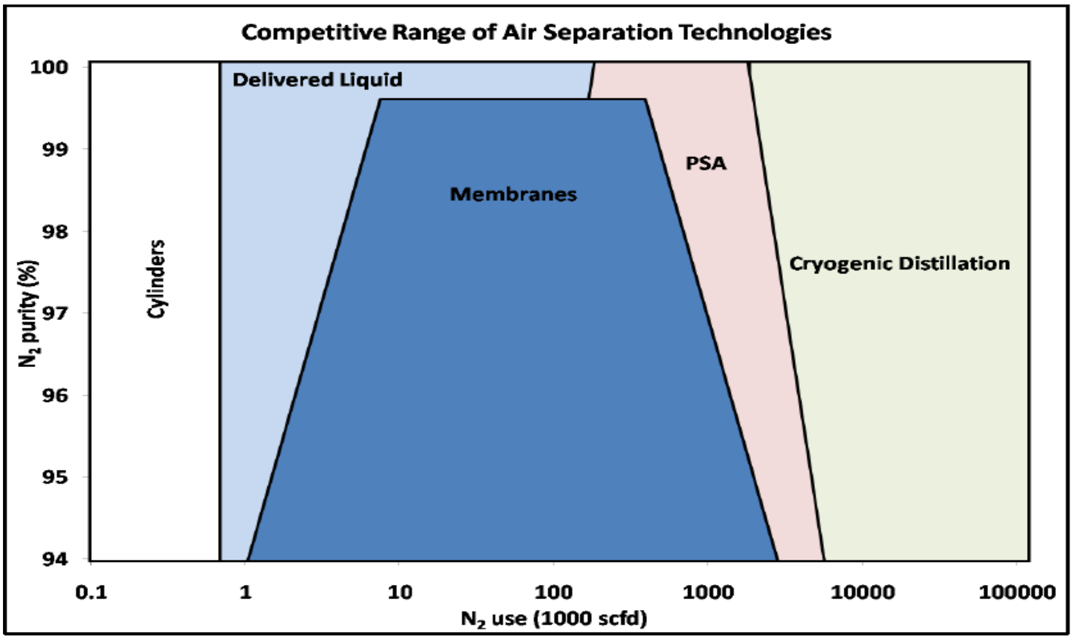


Figure 25 – Competitive ranges of Air Separation Unit – image from [27]

Considering the quantity of ammonia that this study is looking at, only the PSA and Cryogenic distillation would be of interest.

The **Pressure Swing Adsorption** (PSA), dried, cleaned, and compressed air is sent into an adsorber towers filled with a carbon molecular sieve (CMS). During pressurization of the CMS bed, oxygen is preferentially absorbed on the surface of the carbon pellets at high pressure, while nitrogen is retained in the feed gas flow. The nitrogen separation process is stopped before the CMS comes to the adsorption capacity limit and the feed air supply is switched over to the second adsorber tower. Nitrogen can be obtained with a purity of up to 99,995% vol. However, with PSA air separation, the purity of the nitrogen 'isn't as good as with cryogenic ASUs.

The **Cryogenic air separation** is mostly used for medium to large scale production plants. Here, incoming air is filtered, compressed and cooled and impurities like water and CO2 are removed. The purified air passes through a heat exchanger which cools it down to a cryogenic temperature of approximately −185°C. The air is then sent to one or two distillation columns to separate the components (gaseous nitrogen, oxygen, and liquid argon) according to their different boiling temperatures. Pure N2 gas is collected at the top of the column and stored at high level of purity.

Cryogenic air separation is the most plausible purification system into the context of this analysis. Going in more details, the main components of a cryogenic ASU are usually: centrifugal compressor, the floating Heat HX, the radial gas, the towers and drivers [27]. The scheme of a cryogenic air separation unit is shown below Figure 26:

Figure 27

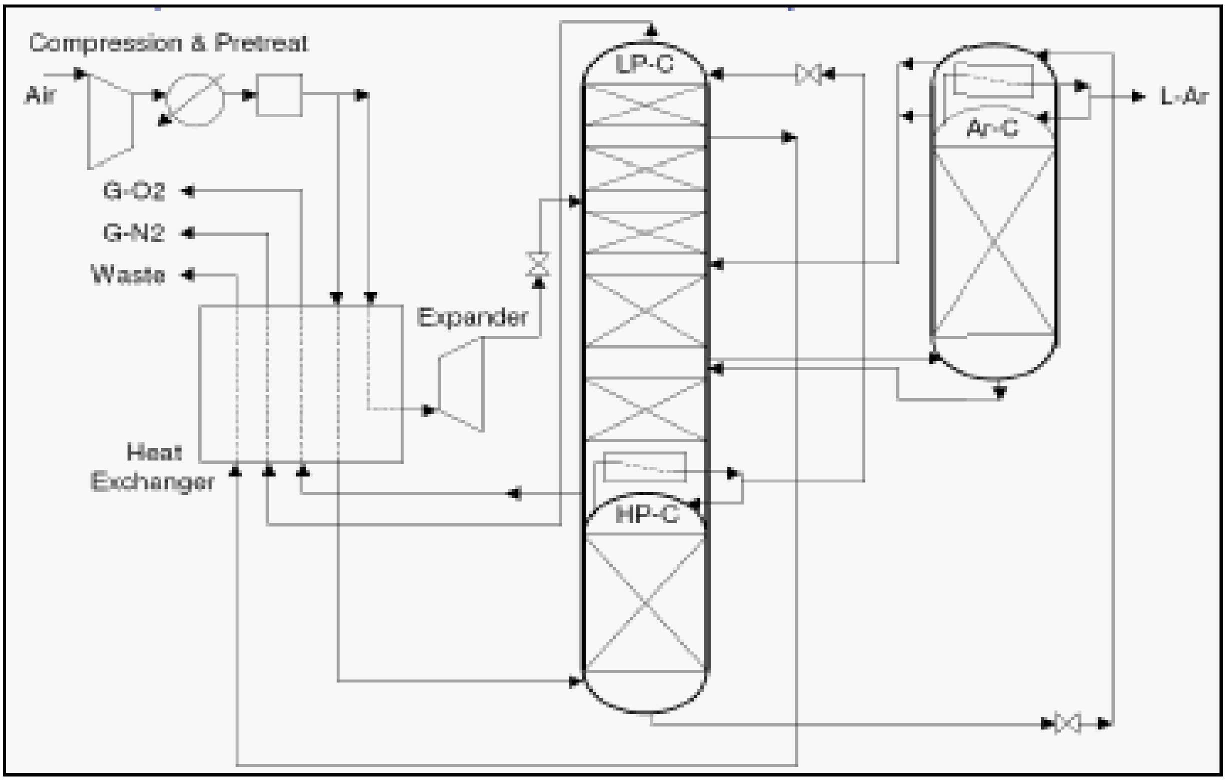


Figure 26 – Cryogenic ASU scheme – from [27]

Understood the technical background is now possible to discuss the costs of ASUs. From a literature review, two cost functions and some punctual estimations have been found, as in Figure 27:

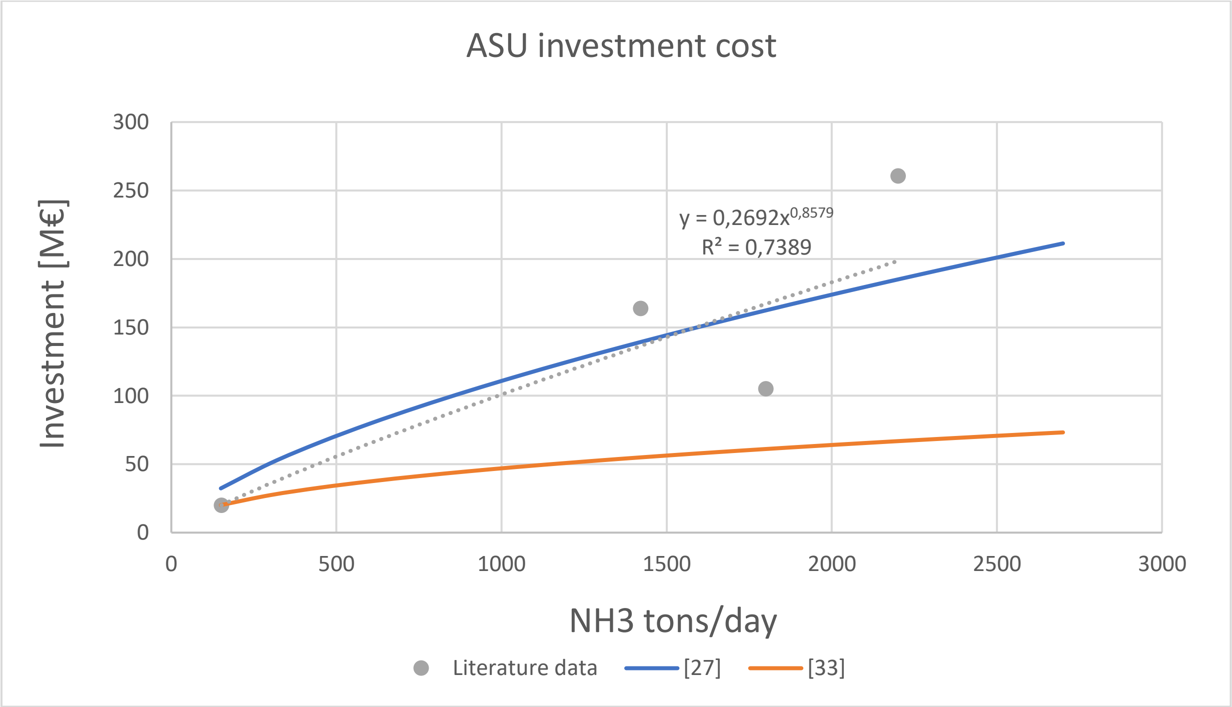


Figure 27 - Air Separation Unit investment cost function

The orange line [33] has a considerably lower investment cost if compared to the blue line [27] and the fitting curve that comes from the data found in literature ([27], [28], [33], [34]). Since such a difference on the cost is not well justified in [33] , we won't consider it in the following calculations. On the other hand, the blue line well fit the fitting curve that comes from the interpolation of the literature data; therefore, it will be used for the evaluation of the levelized cost of ammonia.

Generally, the ASU accounts for 1/3 of the overall ammonia synthesis plant investment, but it can also be a source of some positive income if oxygen, a residual product from the air separation, is sold. Nevertheless, in this study, selling oxygen is not considered as an income.

3.4.2 Haber-Bosch synloop

The Haber-Bosch synloop is named after its inventors, the German chemists Fritz Haber and Carl Bosch, who developed it in the first decade of the 20th century. The process converts atmospheric nitrogen to ammonia (NH_3) by a reaction with Hydrogen (H_2) using a metal catalyst under high temperatures (between 400 and 500 °C) and pressures (above 10 Mpa).

The ammonia conversion, given these extreme conditions, absorbs about 1% of the 'world's total energy production, so its conversion to sustainable production is more than desirable. Moreover, this high energy-intensity process has different impacts on the environment and ecosystems such as the leaching of nitrates into groundwater, rivers, ponds and lakes and causing anthropogenic disruption to the nitrogen cycle.

The HB synloop is the most common technology adopted nowadays, but not the only one. There are some new research fields on the use of absorbent-enhanced synloop, but as can be seen in the following Figure 28, the costs, within the range of our interest, are comparable, so it won´t have any remarkable influence on the cost estimation of the NH3 transport.

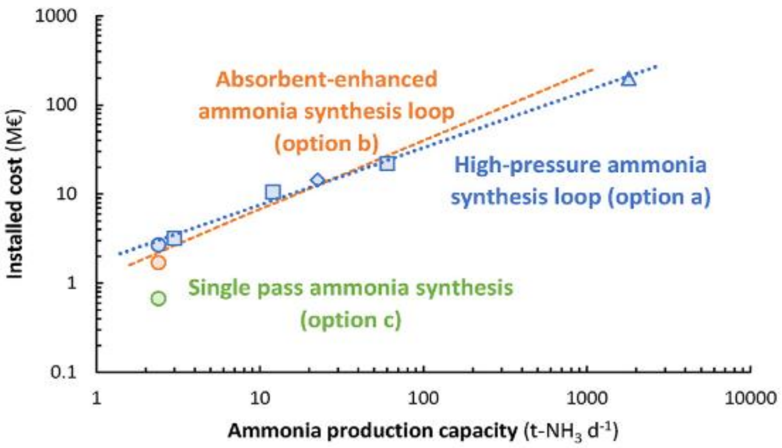


Figure 28 – Ammonia production capacity (tNH3/day) - from [36]

Despite the longevity of this process, it is still a really expensive mechanism to produce ammonia: about 2/3 of the total investment for the ammonia synthesis plant is constituted by the HB synloop.

For the HB cost two cost functions have been identified, one from [27] and another from [33]. In addition to that, also two punctual values from [33] and [34] have been plotted, which make more reliable the approach. All of them are displayed in the following Figure 29:

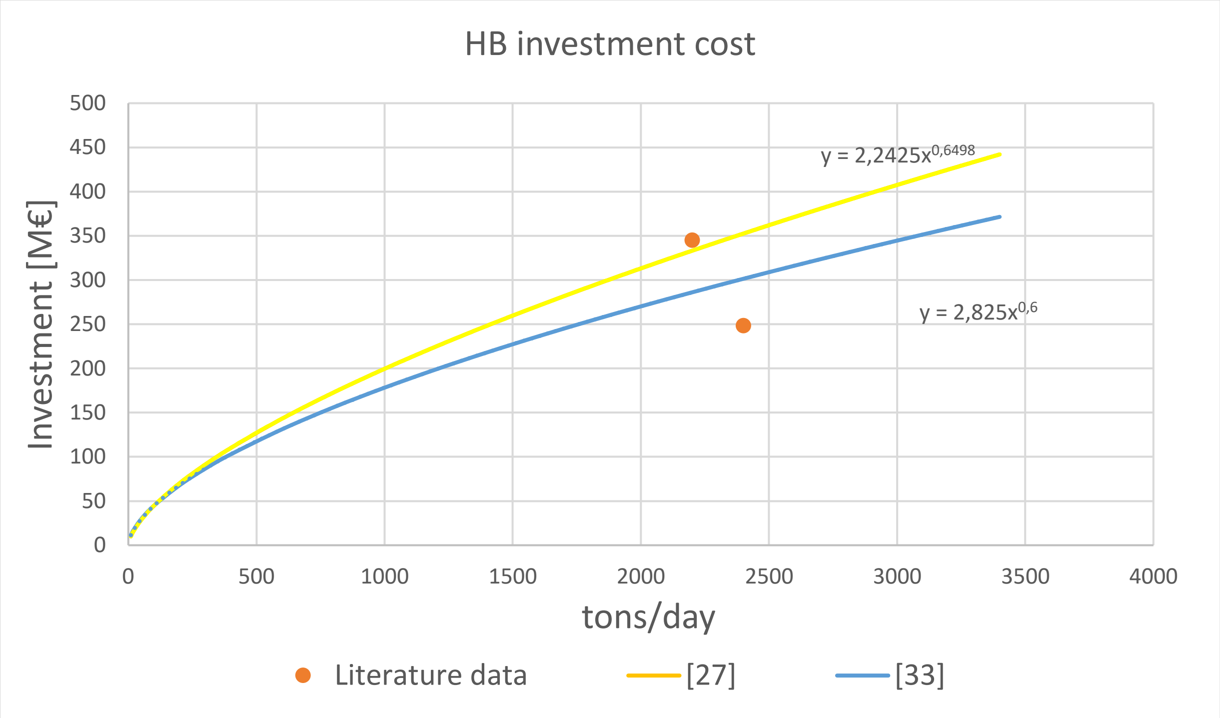


Figure 29 - Haber-Bosch investment cost function

The two cost functions have roughly the same behavior: The blue line with more scaling behavior compared to the yellow line, but overall, the values found for the HB investment cost seem to be consistent among the available bibliography read.

The operational expenditures of a synthesis plant are mainly related to the purchase of the energy needed by the ASU and HB during their daily operation. The synthesis of ammonia is a quite energy-consuming process, reaching up to 15% of the ammonia production cost itself.

For the evaluation of the operational costs of the whole system (ASU + HB), the synthesis plants have been divided in 3 categories: small, medium and high daily capacity. The values used for the evaluation of the LCOA are listed in the following Table 7:

Table 7 - Ammonia synthesis plant energy consumption assumptions

Plant size	**NH3 daily production**	**Energy consumption**	**Reference**
Small	< 50 tons/day	1000 kWh/tNH3	[33]
Medium	50 < tons/day < 800	650 kWh/tNH3	[35]
High	Tons/day > 800	400 kWh/tNH3	[36]

For the CAPEX it has been possible going through two approaches: one according to a bottom-up analysis of the cost and one considering the cost of the system as a whole. Indeed, the cost of the ammonia synthesis plant has already been divided into two main components, the ASU and the HB, but in the literature, they have been found studies that consider the system as a whole and so providing "" the investment cost without any distinctions among the components, as displayed in Figure 30:

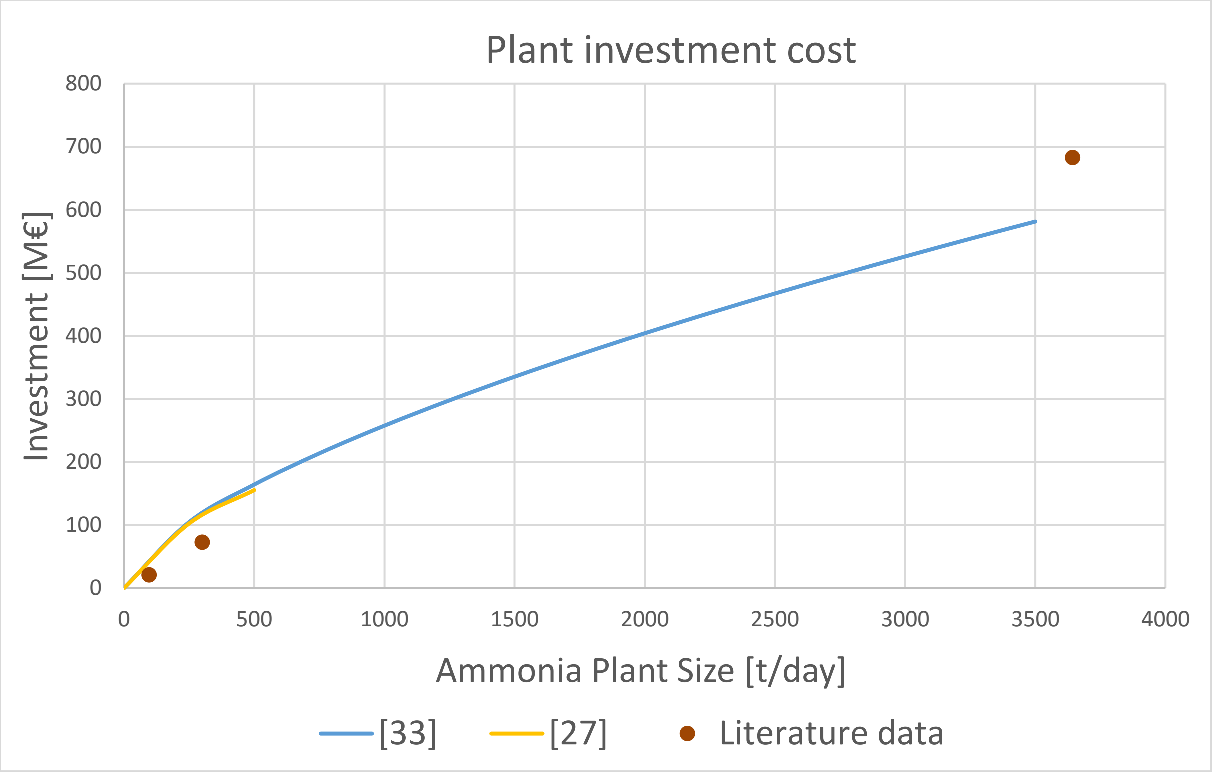


Figure 30 - Ammonia plant invest cost function

For the overall ammonia synthesis system, 2 different cost functions have been found:

[27]: $y = 3{,}73x^{0{,}6}$ Yellow line
[33]: $y = 2{,}8911x^{0{,}6499}$ Blue line

Since the equation from [27] (yellow line) is valid only for small plants between 0 and 150 MW of NH3, it will not be considered for the final cost function estimation, but it is still important to mention when small plants are involved. Besides, some values for singular plants have been found in [27], [35], [37] and are consistent with the trend seen in the cost functions.

Having now the investment cost from both the approaches ready, they can finally be compared to choose then the most reliable approach to be used in the next section" calculation.

Two or more cost functions for the bottom-up analysis could be plotted based on Figure 27 and Figure 30 since, for both the ASU and HB, two cost functions have been highlighted. To make it easier, two-case scenarios are presented: for the best-case scenario (red line), the lower cost function for the HB and ASU have been added, while for the worst-case scenario (Green line) the opposite.

Finally, 2 cost functions from the bottom-up analysis are given and one comes from the literature approach (Blue line) considering the plant as a whole.

These functions are now shown in Figure 31:

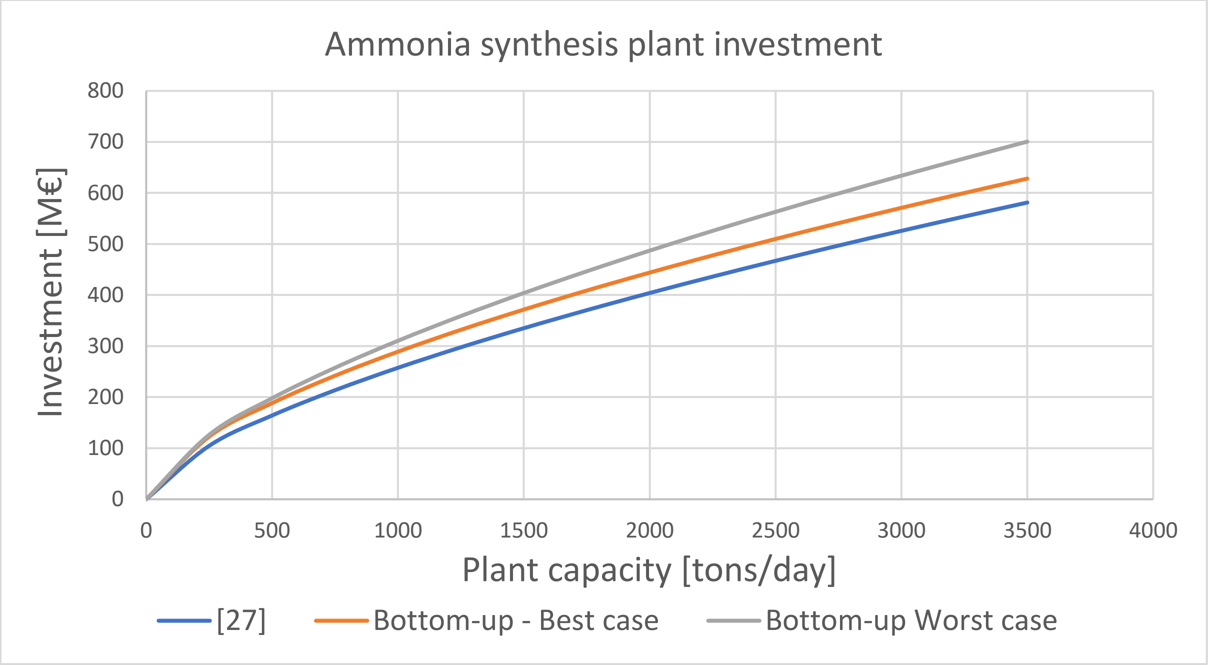


Figure 31 - Ammonia synthesis plant investment cost for different approaches

Especially for small plant capacities, the three cost functions tend to overlap, diverging for higher values of ammonia daily production. Nevertheless, the trend appears to be quite understandable in all the three cases, showing a scaling economy for larger sizes.

Since is not easy to choose one approach or another, being the specific technical and economic characteristics of the system not fully explained in the bibliography, it seemed wise to understand before the weight of the ammonia synthesis plant' 's investment on the global LCOA.
In Figure 32 there are therefore displayed the main features of the system that contribute to define the levelized cost of ammonia. The main cost is obviously the hydrogen production from green electrolysis (70%), while the percentage of the other parameters change slightly depending on the number and dimension of the ships used for the ammonia transport, but this will be discussed in more detail in the next chapters.

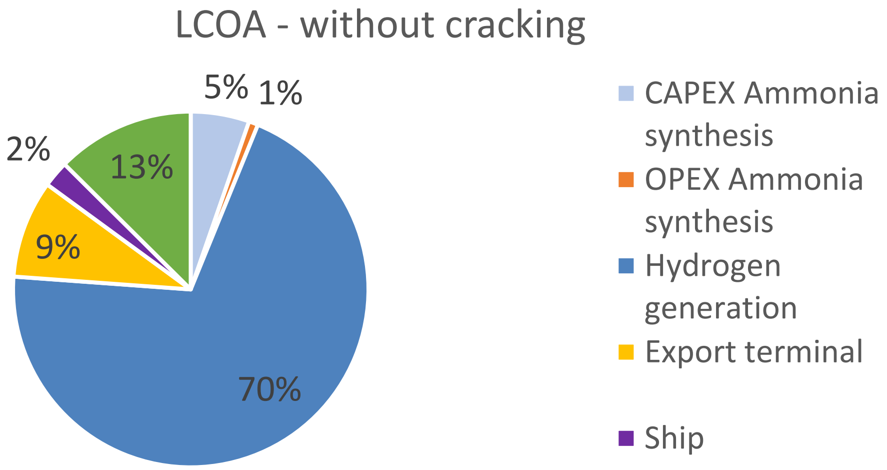


Figure 32 – LCOA breakthrough without cracking – Example for a 3500 tons/day ammonia production plant

Looking at Figure 32, the ammonia synthesis plant has a marginal weight on the overall LCOA, it indeed accounts roughly for just the 5% indeed. Therefore, it seemed reasonable to pick an average value among the 3 cost functions mentioned before, having such a small influence on the euro/kg of the ammonia transport. The final equation is $y = 2{,}7003x^{0{,}6658}$ and its behavior can be seen in Figure 33:

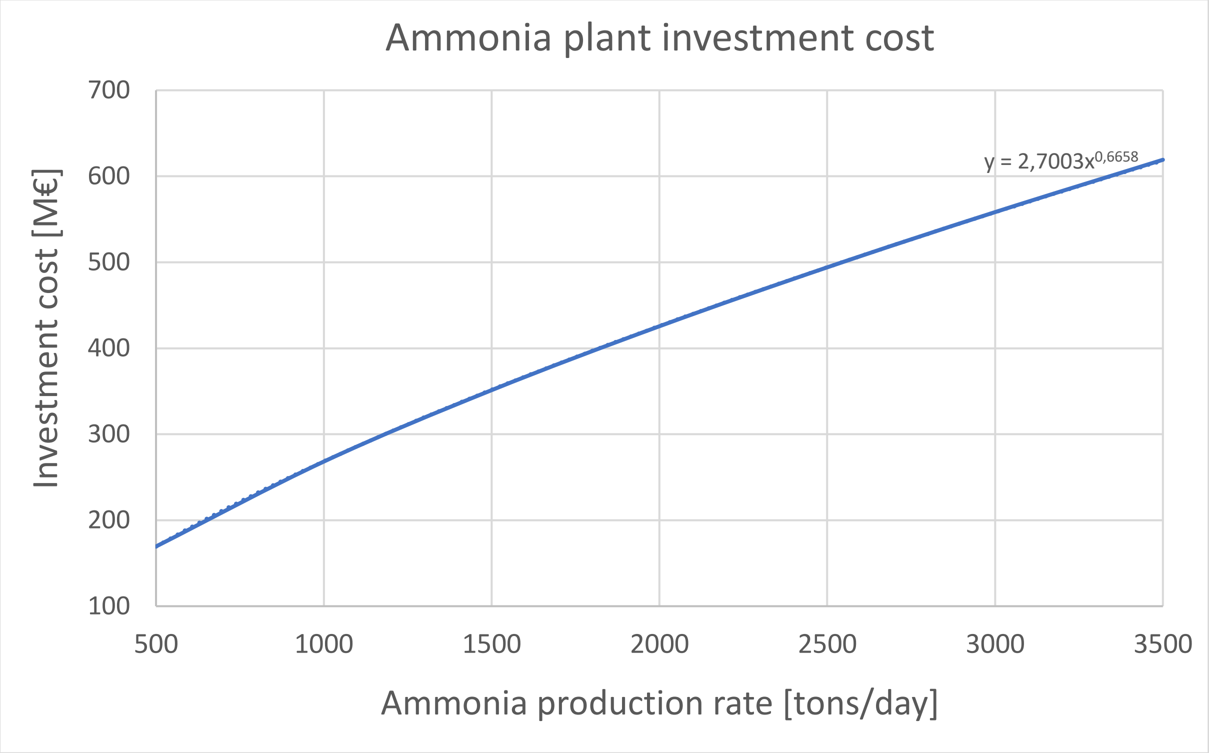


Figure 33 - Final ammonia synthesis plant investment cost function

3.5 Liquid Ammonia Storage

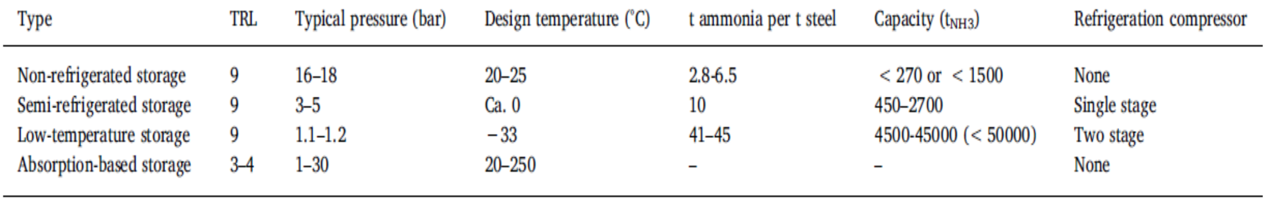
The ammonia storage is gaining significant appeal as a green fuel since the boiling point of liquid hydrogen is much higher (-33°C) if compared to the liquid hydrogen (-253°C); besides this, it also contains more hydrogen per weight than all the other hydrogen carriers (liquid and gaseous H2), and it is an already mature technology across the world because of the fertilizer sector [38].

Depending on the tons of ammonia stored, different technologies can be adopted varying from non-refrigerating and pressurized storage to refrigerating storage (with liquid ammonia) to a combination of them. Pressurized not-refrigerated storage at ambient temperature are most used for small capacities (till 270 tons): the typical pressure is close to 17 bars and the energy density at this state is 13,77 MJ/l [27].

For larger quantities, the ammonia needs to be liquified to lower the costs of the tank and the footprint. Semi-refrigerated storages are adopted for small capacities (till 270 tons), undergoing low pressures (3-5 bars) and temperatures close to 0°C. For uses like trade and transport, where bigger tanks are involved, the ammonia is stored in the liquid form at ambient pressure with a temperature of -33°C and an energy density of 15,37 MJ/l [27].

The general characteristics of the storages mentioned before are listed in Table 8

Table 8 - Ammonia storage typical features and values – from [38]



Type	TRL	Typical pressure (bar)	Design temperature (°C)	t ammonia per t steel	Capacity (t_{NH3})	Refrigeration compressor
Non-refrigerated storage	9	16–18	20–25	2.8-6.5	< 270 or < 1500	None
Semi-refrigerated storage	9	3–5	Ca. 0	10	450–2700	Single stage
Low-temperature storage	9	1.1–1.2	− 33	41–45	4500-45000 (< 50000)	Two stage
Absorption-based storage	3-4	1–30	20–250	–	–	None

Depending on the technology adopted the ratio of the tons of ammonia contained over the tons of steel used for the storage changes significantly. Pressurized storage, since they have to bear the pressure inside the tank, can store between 2,8 and 6,5 tons of ammonia per ton of steel, while for low-temperature storage (Liquid Ammonia), this ratio goes up to 41-45 because the tank must bear just the static pressure [27], [33].

Nevertheless, low-temperature storage suffers from Boil Of Gas (BOG) losses as it happens for liquid Hydrogen, even though the percentage of daily losses is lower because the boiling point is closer to the ambient temperature in this case. The BOG must be handled, and this requires an energy input to keep cooling down the carrier. This does not happen for the pressurized storage instead: once the tank is pressurized, no more energy is needed.

Ammonia can be stored also in other ways, even though not yet mature, like absorption storage materials [38]. In this case a higher degree of safety is reached because ammonia can be desorbed only supplying heat or increasing the pressure. With this technology, 5-10% in weight of ammonia can be reached.

3.5.1 Materials

Only low-temperature storage will be considered in this analysis, given the high volumes. A scheme of this kind of storage is shown in Figure 34.

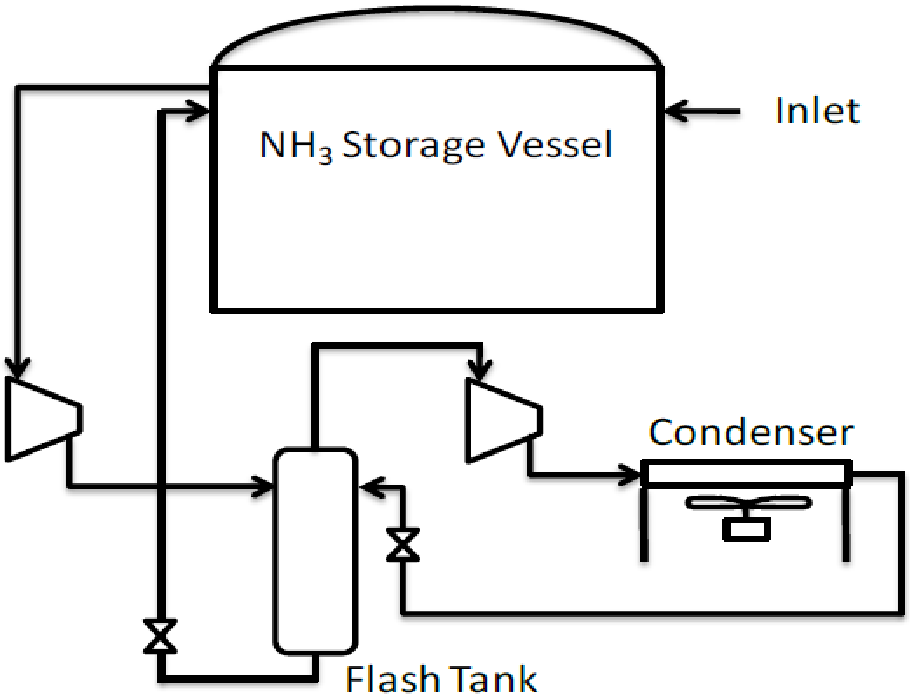


Figure 34 – Ammonia storage technical scheme

Liquid ammonia lies in the main tank, BOG is collected and compressed before flowing through the flash tank to be cooled down. After that, another compression occurs to a value above the condensation pressure. Once the ammonia is separated, it enters the flash tank, expands, and separates into liquid and vapor form. At this stage, the vapor goes again through the second stage compressor to start the cycle again. The reliquefied flow at the bottom of the flash tank is sent back to the storage after being expanded to the tank pressure through a valve [33].

There are two types of tank construction for low-temperature storage: single and double walled [38]. The double-walled configuration is more expensive than the single because of the insulation that fills the gap between the two walls, significantly reducing the BOG rate. The double-walled configuration has higher safety because the two containment layers are more reliable in case of cracks develop and lower maintenance and operating costs.

The trade-off between the two technologies is based on the time of the storage itself. For a longer period of storage, the double walled storage could limit the losses for BOG and the energy requirement for the cooling. In contrast, the single single-walled can be the best option for storing ammonia for a limited period, given the relatively low investment cost. The key features of the double walled storage are reported below in Table 9, with a comparison with the LNG storage.

Table 9 - Ammonia storage materials

Parts	LNG	Refrigerated Ammonia
Primary liquid container	9-percent nickel (Ni) steel	Low temperature carbon steel
Secondary liquid container	Dike	Dike
Suspended deck	Aluminum	Low temperature carbon steel
Vapor container	Carbon steel	Carbon steel
Bottom insulation	Cellular glass blocks	Cellular glass blocks
Side insulation	Resilient blanket and perlite powder	Resilient blanket and perlite powder
Suspended deck insulation	Fiberglass blanket	Fiberglass blanket
Pump well and internal piping	Stainless steel	Stainless steel or low temperature carbon steel

The energy consumption required for the liquefaction process and the cooling of storage is much lower compared to the energy consumption of the liquid Hydrogen: the liquefaction of Hydrogen required a huge amount of energy (10-12 kWh/kgH2), being the condensation temperature below 20 Kelvin. For the ammonia liquefaction process the energy is roughly 0,033 kWh for the equivalent of 1 kgH2, almost 3 orders of magnitude lower, even because the outlet temperature of the ammonia synthesis is around -25°C, really close to the -33°C needed to liquify NH3 [33]. The energies required for the cooling of the storage itself are also sensitively different even though less important than for the liquefaction process: 1,812 kWh/kgLH2 vs 0,181 kWh/kgH2eq, [33]. The cost assessment of the storage, as it has been done for all the other important features of the ammonia production system, has been carried out through a detailed literature research; the results of this analysis are shown as follow in Figure 35:

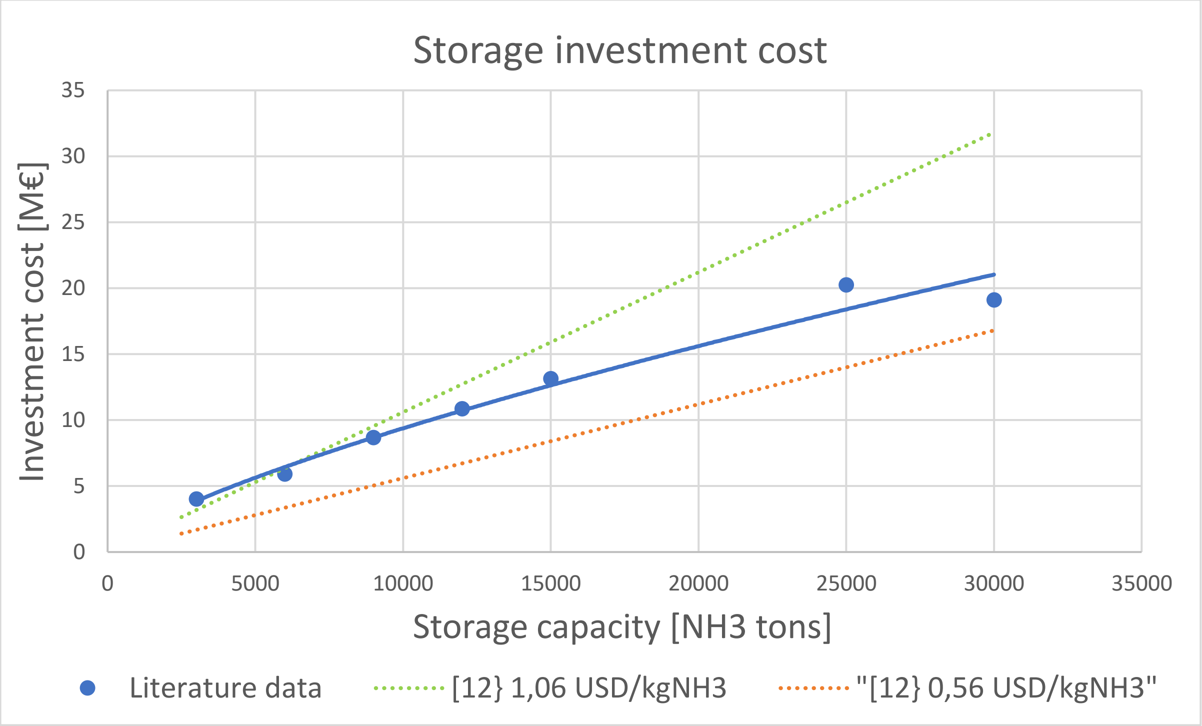


Figure 35 - Ammonia storage investment cost

The blue dots represents punctual estimations from different studies ([27], [39], [40]) and the blue line it is the resulting fitting curve from their interpolation. The green and the orange line are instead the range that [38] suggests as upper and lower limit for the specific cost of the liquid ammonia storage: between 1,06 USD/kgNH3 (green line) and 0,56 USD/kgNH3 (orange line), including the construction, surrounding bunks and refrigeration loop (as shown in the previous storage system scheme). The dots found in other sources fit well these assumptions, close to the upper limit for low capacities and close to the lower limit for high capacities, giving to this approach a high degree of reliability. The fitting curve has equation $y = 0{,}0068x^{0{,}7877}$ where ""x"" is the capacity in tons of NH3 of the storage, and it will be adopted as cost function for the storage investment.

It is worth to mention that the biggest tanks that are available nowadays worldwide do not exceed 45 000-50 000 tons of ammonia (50000-55000 m3) [41], [27], therefore for high capacities more than one storage tank must be considered. Making a comparison with the liquid hydrogen storage tank, the behavior of the cost function is quite different: in the last chapter it has been shown that the LH2 tank doesn't have a remarkable scaling behavior, on the contrary, it is quite straight instead, with a scale-up coefficient approximately equal to 0.85. The same behavior has been searched for the liquid ammonia storage, but as it is shown in Figure 36, the cost function that comes from the interpolation of literature data overlaps with a curve that has a scale up factor equal to 0,7.

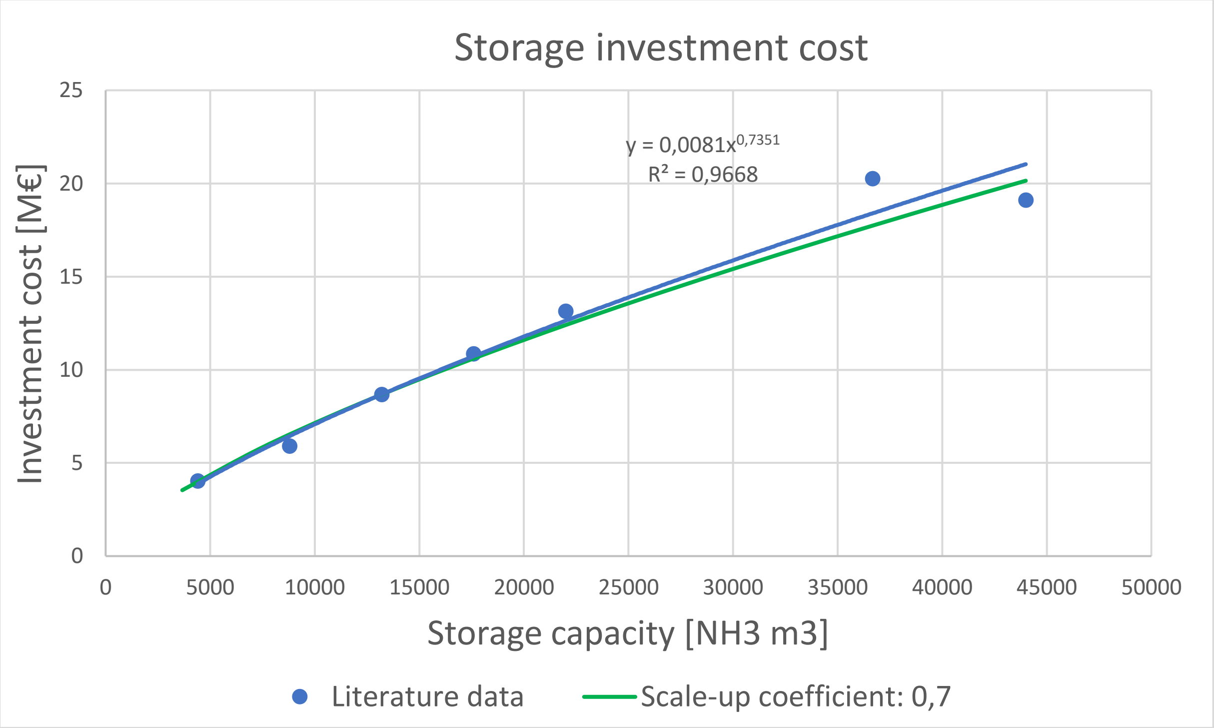


Figure 36 - Ammonia storage investment cost - scale up approach comparison

Such a low coefficient is a sign that in the ammonia industry there is still an economy of scale, probably due to the well-established techniques in this field. The not-scaling behavior of the liquid hydrogen storage led us to make some assumptions, which in this case cannot be done. Nevertheless, this information will be very important for some reasonings that are going to be commented on the next sections.

3.6 Liquid Ammonia Tanker

As it happened for in-land storage tank, also the liquid ammonia ships can fall into three categories depending on the ammonia capacity that they carry: fully pressurized, with independent Type "'C' tanks; semi-refrigerated, with independent Type "'B' tanks; and fully refrigerated, with independent Type "'A' tanks.

The logic is quite the same seen before: fully pressurized vessels are used to deliver ammonia over short distances at pressures up to 20 bar, but due to the large pressures their tanks are very heavy, so they are suitable only for small capacities (below 4000 cubic meters). Semi-refrigerated vessels capacities can lie between 1500 and 30,000 cubic meters. Fully refrigerated vessels can have 15,000 to 85,000 m3 of capacity and are used to transport ammonia over long distances by sea. These Gas carriers transport Liquid ammonia at -33°C with an onboard refrigeration system working in the same way as a land-based refrigeration storage [38].

The majority of NH3 is traded for long distances via two types of vessels: Medium Gas Carriers (MGCs) and Large Gas Carriers (LGCs), both adopting independent type 'A" tanks. At the current state of the art, almost all the ammonia production is delivered by MGCs. For high quantity of NH3 to deliver, Very Large Gas Carrier can be adopted, usually they can also serve for LPG transport. Nevertheless, at this stage there is still not a widely spread because the ammonia terminal facilities have usually a limited capacity [38].

Even for the evaluation of the NH3 tanker it has been conducted an analysis of the available literature and the results are now shown in Figure 37:

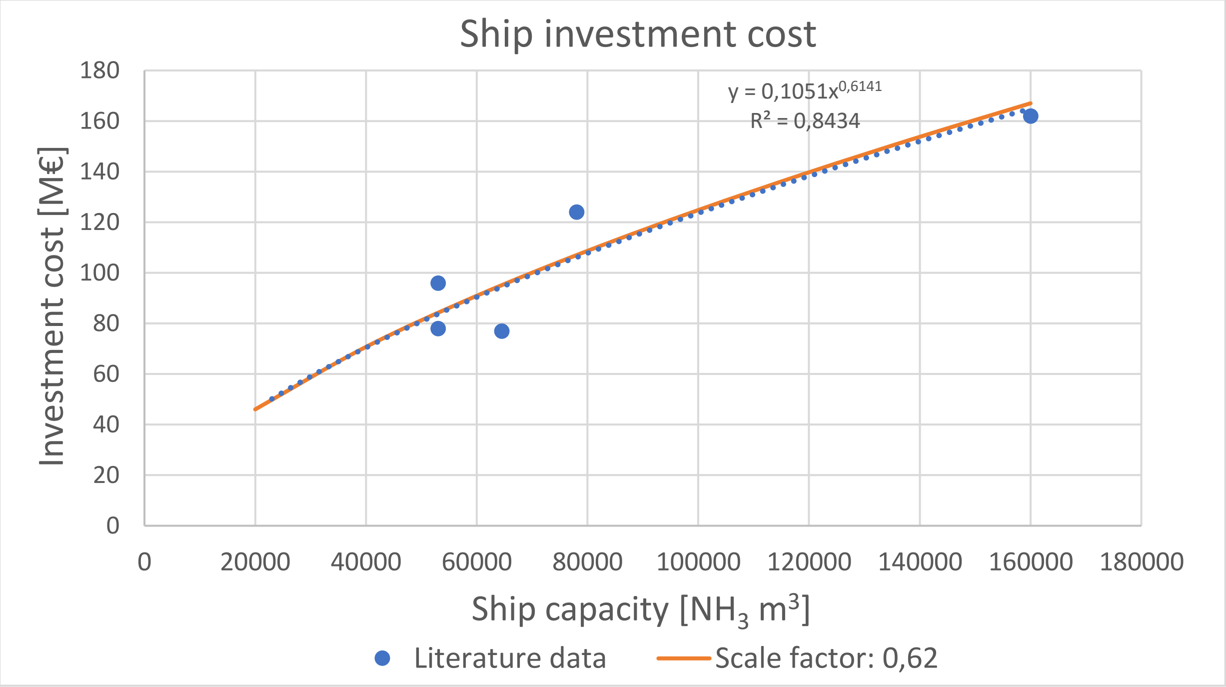


Figure 37 - Ship investment cost function and comparison to scale-up approach

The dots come from [14], [23], [7], [42], [43] and the resulting fitting curve has an equation equal to: $y = 0,1051x^{0,6141}$.

On the same graph is possible to notice how the fitting curve overlap with a cost function that has a scale factor equal to 0,62. This is quite consistent with what has been found with the storage cost function, which had a scale up coefficient of 0,7. Therefore, unlike what happened for the liquid hydrogen case, where different behaviors for storage and ship were plotted, here the data tend to be coherent; a further bottom-up analysis is then excluded from this work.
It is also worth to mention how the specific cost of the tanker, in €/m3 of NH3, varies regarding its capacity. In addition to the sources already mentioned, [38] and [42] give a specific cost for the tanker. All these values are shown in Figure 38, and for high capacities (more than 120000 cubic meter) they all tend to overlap (blue dots, curve and dashed lines), strengthening the approach for bigger vessels.

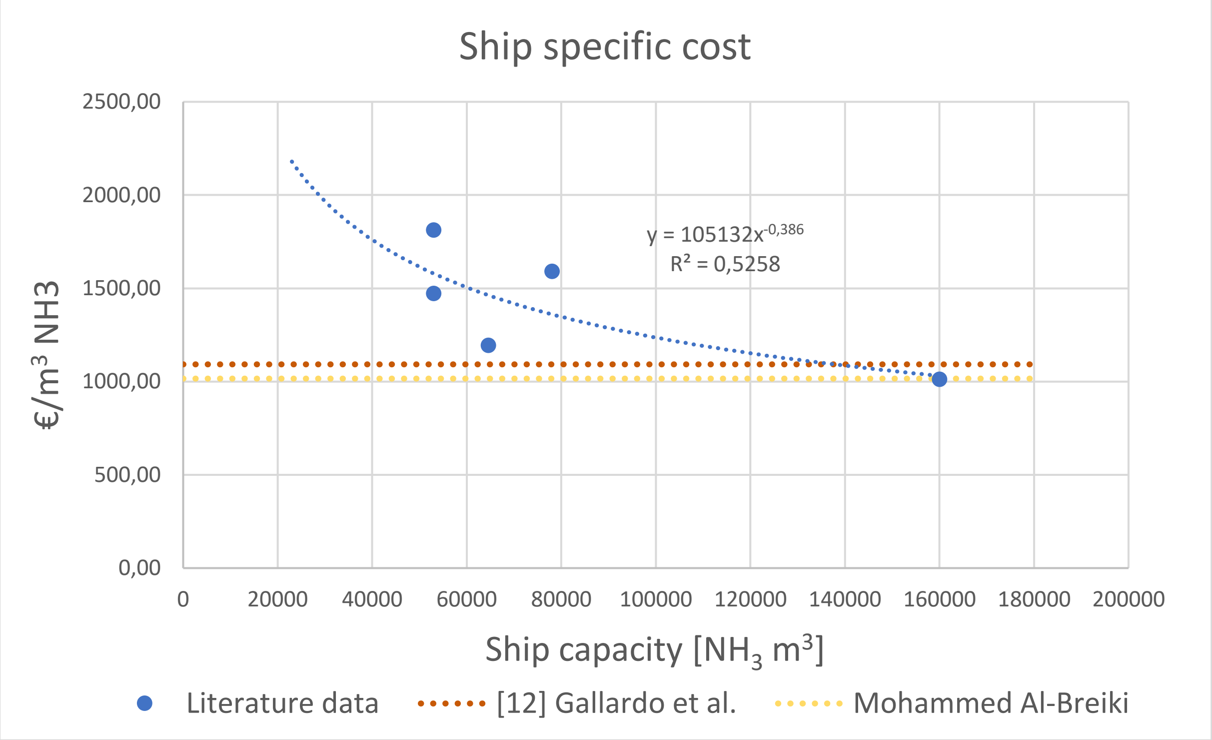


Figure 38 - Ship specific cost

On the OPEX side, in all the reference mentioned before there is general agreement on the fuel consumption (2500 MJ/km), which include the energy consumption of re-liquefaction of BOG onboard the NH3 carriers as well. The BOG is fixed to 0,2% per day of traveling [43], lower than the LH2 ship (0,5%/day), but higher than the BOG that takes place inside the storage that are utilized in land.
The choice of one ship is not mandatory. The project can involve different assumptions and the best choice could be to construct smaller terminals and several smaller ships, if the annual ammonia delivered is fixed.
From Figure 37 and Figure 38 it can be noted that the investment cost for an ammonia tanker is not linear, and for bigger ships the cost tends to be lower. Therefore, assuming the overall quantity of ammonia delivered during the year fixed, increasing the number of the ships means increasing the cost of the total fleet. This is clearly showed in Appendix II.

3.7 Import and Export Terminals

The ammonia terminals, both export and import, do not differ too much from the design of a liquid hydrogen terminal [43], [7]. The main features of the systems are:

- Storage tank/tanks
- Jetty
- Loading arms
- Pumps
- Piping, buildings, valves, instrumentation
- Safety equipment
- BOG handling system

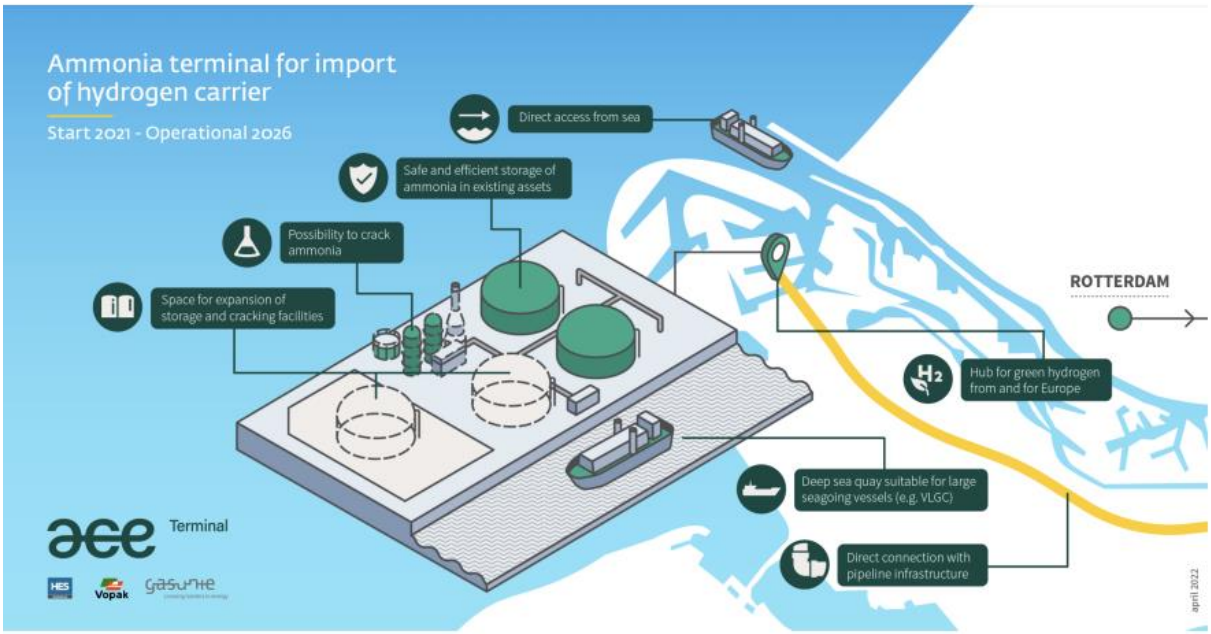


Figure 39 - Ammonia import terminal example - image from "Ammonia Energy Association"

The main difference between an import and export terminal consists of the capacity of the terminal itself. Usually, for import terminals the capacity is higher since it could receive the energy carrier from different exporting terminals. The cost difference between export and import terminals is then allocated mainly on the storage cost.

Depending on the desired final energy carrier, NH3 or H2, it can also happen that the import terminal has a cracking facility, like in Figure 39, which constitutes a relevant part of the terminal CAPEX and OPEX. In this work it will be studied separately in the next section, since the cracking facility is not always part of the export terminal, and the ammonia can also be sold in its molecular form, avoiding further conversions.

It has been already pointed out in the past sections that, compared to the liquid hydrogen study case, both the storage and the ammonia tanker follow economies of scale, understandable from their cost functions curves; indeed, they have comparable scale factors, 0,7 for the storage and 0,62 for the ship.

Moreover, the storage cost "s weight over the terminal investment falls always in a range between 40 and 45% Based on this, since the most important cost driver of the terminal follows an economy of scale (tank), it has been decided to proceed with the literature scalability approach (equation (2. 2)) to evaluate the investment of the terminal:

$$C(S_1) = C_0 \left(\frac{S_1}{S_0}\right)^n$$

With:

- C_0 = 76,2 M€ [17], [7]
- S_0 = 34100 tons [17], [7]
- n = 0,67 [43]

The equation (2. 2) indicates an economy of scale approach for the cost evaluation of a given system: the lower the scale-up coefficient the more the economic gain for bigger capacities, since the cost function tends to get less steep as can be seen in Figure 40, where two different scale-up coefficient are compared.

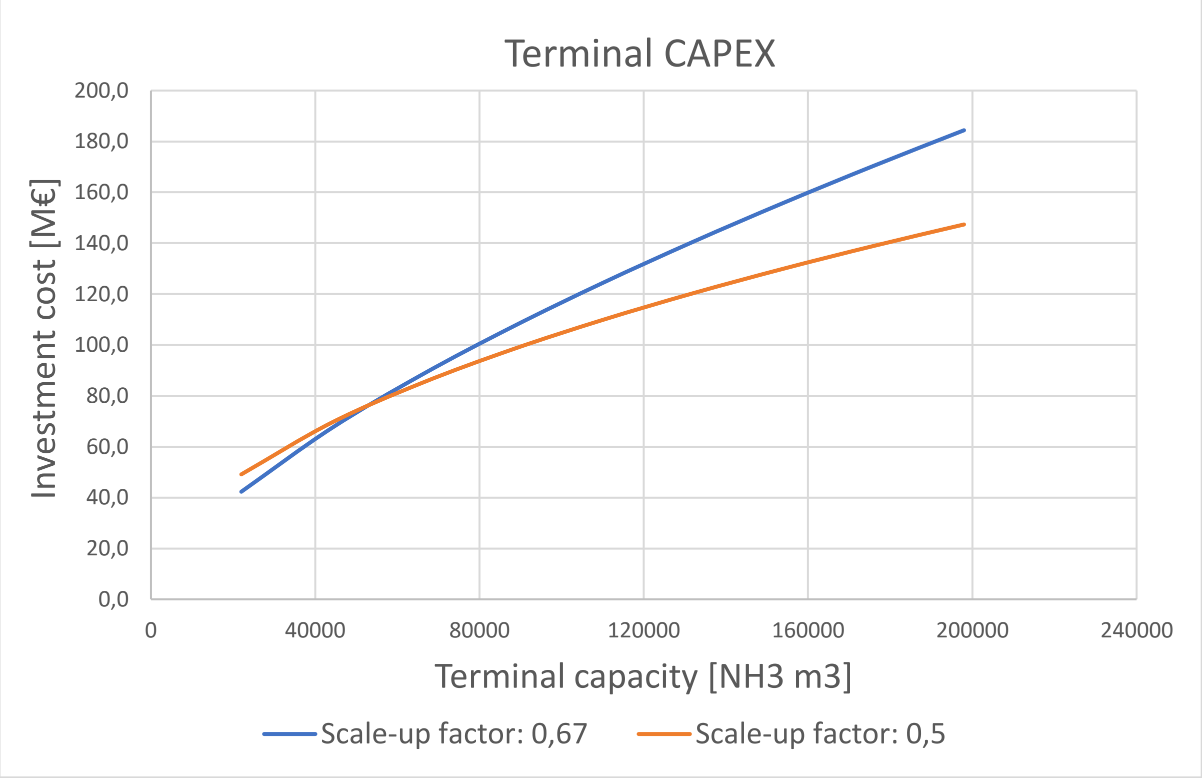


Figure 40 - Terminal CAPEX as a function of its capacity - scale-up factor comparison

The capacity of the terminalis then the only cost driver, and its value depends on 2 factors:

1. The daily ammonia production (tons/day)
2. The number of ships involved in the scenario

While the consequences on the terminal capacity of increasing or decreasing the daily ammonia production are easily predictable, it is worth spending some words on the number of ships involved. Fixed the daily and annual production of ammonia, increasing the number of ships means increasing the frequency of loading and unloading ammonia from and to the terminal. The storage capacity on both the ships and terminals has then to be sized for lower quantities of ammonia, since the loading terminal had less time to accumulate NH_3. On the other hand, choosing a larger fleet also means higher fuel consumptions, higher crew members and overall ships' maintenance. The trade-off of all these cost-drivers will be discuss in the LCOA section.

The effects on the total terminal investment with respect to the number of ships, for different value of ammonia production rate, is shown in Appendix II.

All the assumptions and hypotheses made for the terminals are listed in Table 10.

Table 10 - Terminals hypotheses

Hypothesis	Value	Unit
Safety oversize capacity factor	1,3	-
Number of NH3 tanks per terminal	4	-
Terminals electricity consumption	80 [7]	kWh/ton_{NH3}
Electricity Price at the exporting terminal	18 [7]	€/MWh
BOG	0,03 [7]	%/day
Scale-up factor	0,67 [7]	-
O&M	4 [7]	% of the CAPEX

3.8 Ammonia Cracking Facility

The cracking of NH3 is the reverse reaction of the NH3 synthesis, and it is an endothermic reaction with Hydrogen and nitrogen as products. There are already few examples of it worldwide, but just small-scale NH3 crackers are available in the metallurgic industry, while large-scale crackers are not mature yet [7]. Being the reverse reaction of the ammonia synthesis, the cracking process is also highly demanding from the energy point of view; additionally, also a purification system has to be implemented, and this equipment for large quantities and high purity of Hydrogen can sensibly raise the overall costs. From [7] it is reported in Figure 41, the mass balance of a hypothetical cracking unit including purification

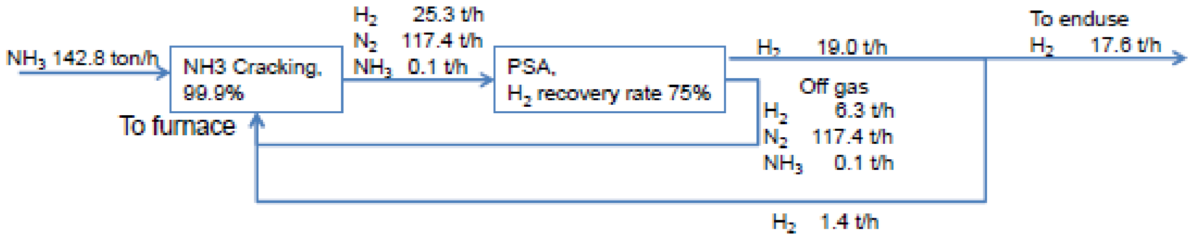


Figure 41 - Example of mass balance for a cracking unit – image from [7]

For a hypothetical flow of ammonia equal to 142,8 tons/h, the output of Hydrogen downstream would be equal to 17,6 tons/h. The discrepancy in the common weight ratio of Hydrogen on ammonia (17,8%) is due to the mass efficiency of both the cracking and purification units.

Being a quite new topic for the industry, not much literature is available on the cost estimation of the Cracking unit. Nevertheless, [7] suggests adopting a scaling approach in this case, which this study will use as well. The result cost function obtained using again equation (2. 2) is displayed as follows in Figure 42.

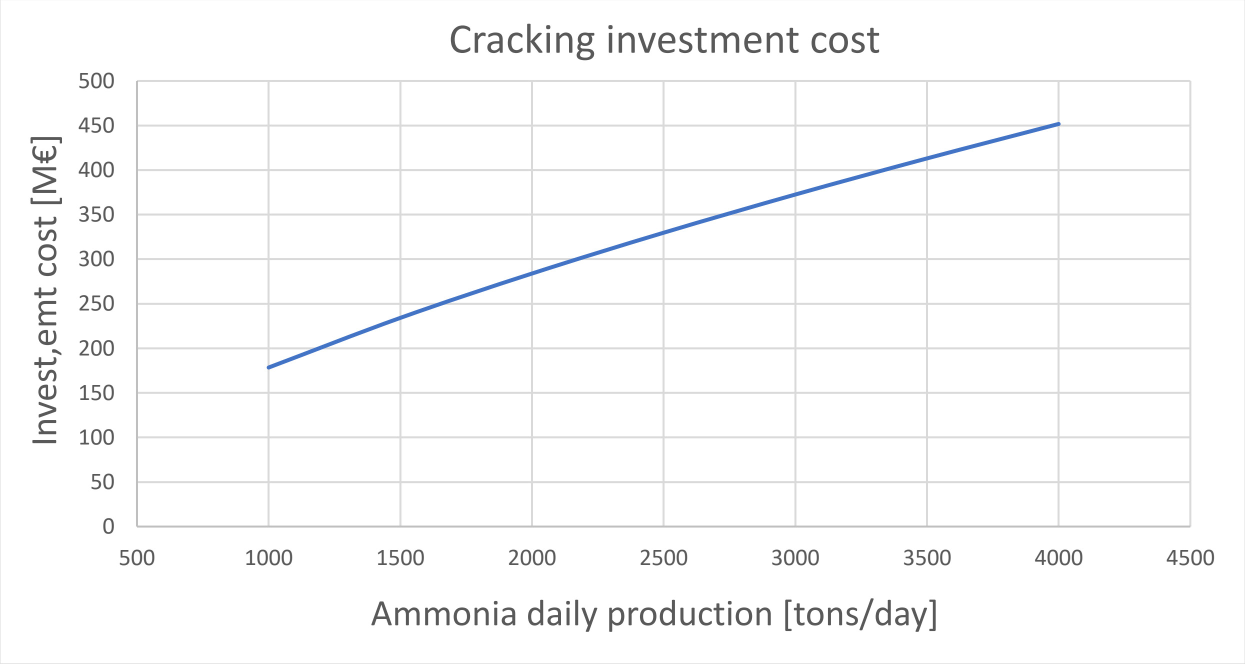


Figure 42 - Cracking facility investment cost depending on the plant size

The scale-up factor is equal to the terminals cost evaluation (0,67), as suggested by [7] while the reference price and size are reported in Table 11 and are taken from [23]. In the same table are listed the hypotheses made to assess the OPEX and O&M of the cracking unit.

Table 11 - Cracking facility hypotheses

Hypothesis	**Value**	**Unit**
Reference cost C_0	460 [23]	M€
Reference size S_0	1500000 [23]	$tons_{NH3}$/year
Scale-up factor	0,67 [7]	-
Specific heat requirement	9,7 [23]	kWh/kg_{H2}
Specific electricity requirement	1,2 [23]	kWh/kg_{H2}
O&M	4 [14], [23], [7]	% of the CAPEX

3.9 Levelized Cost Of Ammonia (LCOA)

The Levelized Cost Of Ammonia (LCOA), as it happened for the LH_2, is evaluated by dividing the sum of the annualized CAPEX and yearly OPEX by the annual ammonia production traded. The yearly NH_3 delivered is equal to the product of the daily ammonia production for the operating days of the system minus the BOG losses of the supply chain (loading, shipping, unloading and storage). As a conservative assumption, the NH3 in gas form that reaches the import terminal is considered a loss, not involving the possible positive income coming from the selling of it. The mathematic process to get the LCOA is shown in equation (3. 1) and (3. 2).

$$LCOA = \frac{\sum_i \frac{CAPEX_i}{a.f._i} + OPEX_i + \; O\&M_i}{NH_3 \; treated \; annually}$$

(3. 1)

Where the actualization factor $a.f.$ is equal to:

$$a.f. = \frac{\left(1 - \left(\frac{1}{\left(1 + \frac{r}{100}\right)^{l.t.}}\right)\right)}{\left(\frac{r}{100}\right)}$$

(3. 2)

With:

- r = discount rate
- l.t. = lifetime [years]
- NH_3 treated annually = [kg/year]

Based on equation (3. 1) and (3. 2) the LCOA has been evaluated and then displayed in a pie chart that shows the relative weights of the ammonia system´s main features. Figure 43 does not account for the cracking related costs, while Figure 44 does.

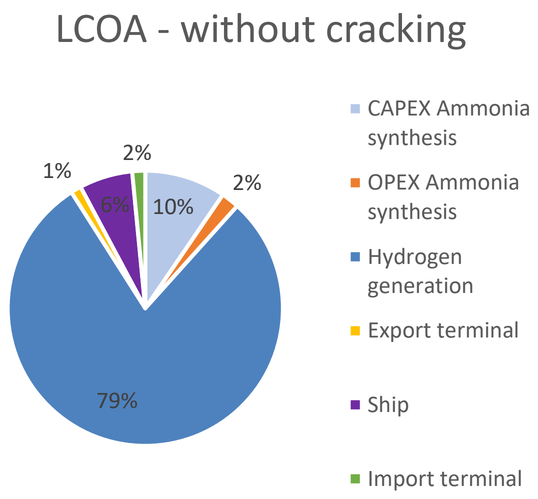


Figure 43 - LCOA without cracking costs

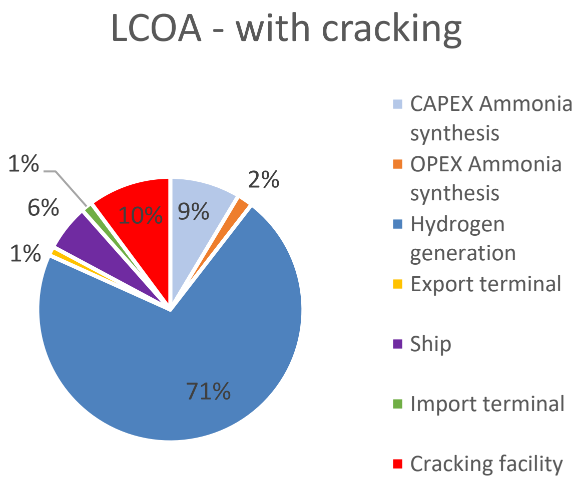


Figure 44 - LCOA with cracking costs

3.9.1 Sensitivity analysis

In this section a sensitivity analysis of the main parameters that affect the LCOA or LCOH will be carried out. But first of all, we need to find a way to compare the costs related to the ammonia transport, to the costs found for the LH_2. To do so and assess the best way of transport, the LCOA has to be converted into a LCOH*[1]; this can be done by means of equation (2. 3):

$$LCOH = \frac{LCOA}{w_r}$$

(2. 3)

Where w_r is the weight ratio of H_2 over a molecule of NH3, equal to 17,8%.

Since now on, all the data reported are computed according to equation (2. 3).

The main variables that affect the LCOH are certainly the distance of transport and the daily ammonia production. In Figure 45 both of them are considered for the sensitivity analysis.

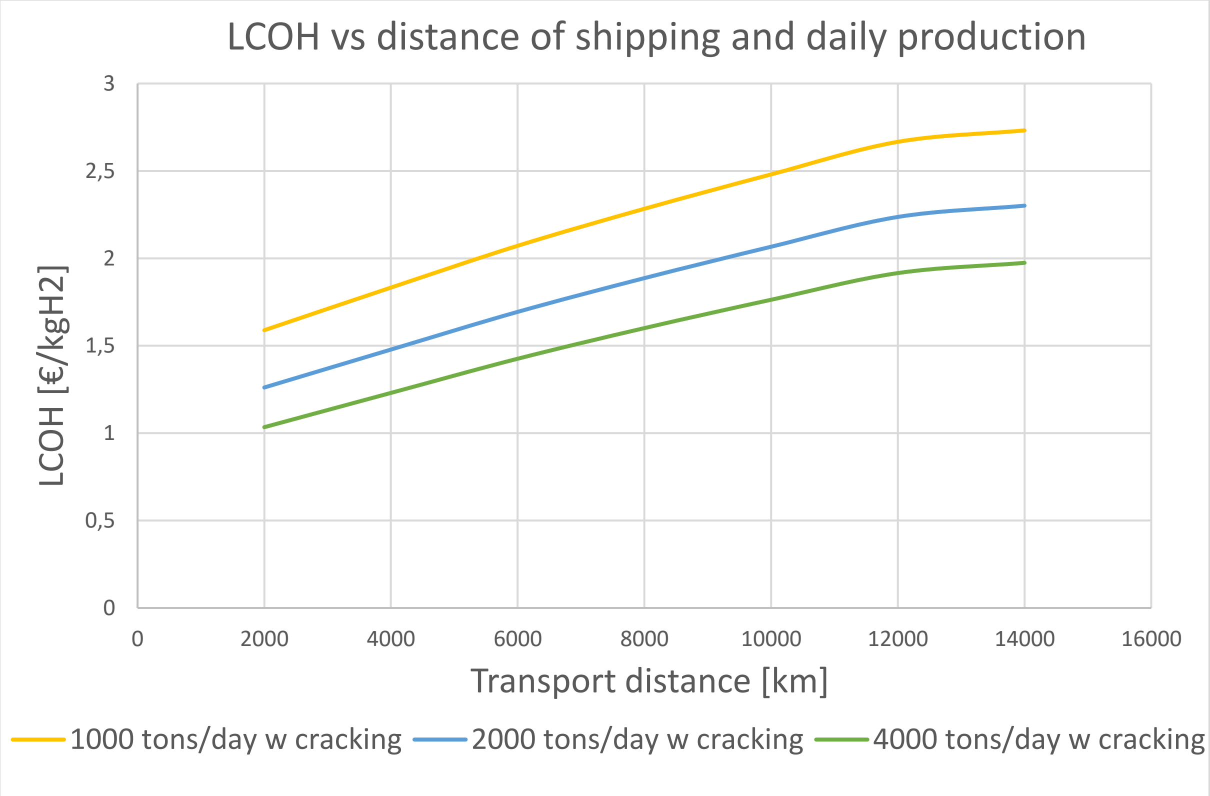


Figure 45 - LCOH vs distance of shipping and daily production

For a given rate of production, the difference between a transport distance of 2000 km and 10000 km is in the range of 1 €/kgH_2, meaning that the transport has a massive impact on the final value. The production, going from 1000 tons/day (violet line) to 4000 tons/day (orange line) has a huge impact as well: for the shortest distance considered (2000 km) there is a difference of more than 0,5 €/kgH_2.

---

[1] *The values that are shown for the LCOH are referring only for the transport chain (Terminals, ship and cracking). The cost of the hydrogen production is not included.

In Appendix II, it is possible to see also the costs without considering the ammonia cracking.

Another important cost driver for the LCOH is the size of the terminal, which is mainly dependent from the ammonia production and number of ships, as already explained in the section dedicated to the terminals´ analysis, Figure 46 addresses this topic.

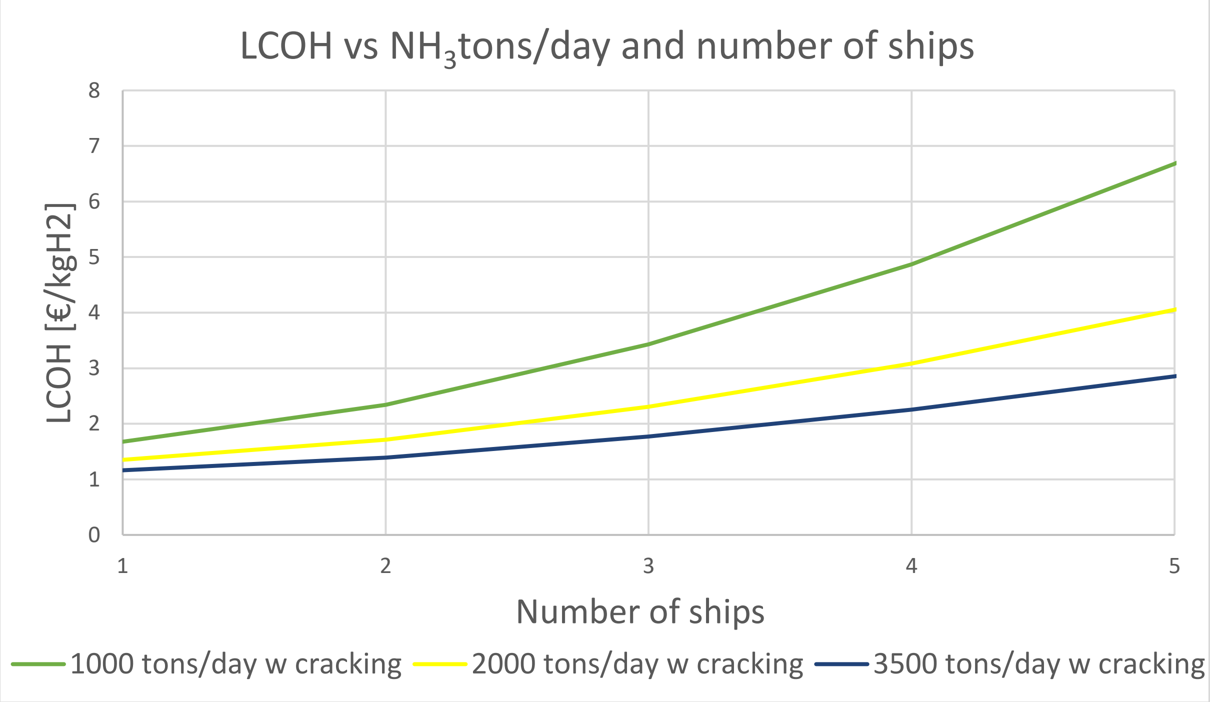


Figure 46 - LCOH vs NH3 tons/day and number of ships – 10000 km case

As already saw in Figure 45, the production is important for the decision making of the project. What is instead interesting to highlight, is that the optimal number of ships is not unique: for a daily production of 1000 tons/day (orange line) the best configuration fall to just one ship, while for a daily production of 3500 tons/day, the best option would be to choose 2 ships.

This trend finds explanation on the different parameters that affects the LCOH changing the number of the ships. Increasing the number of the ships, the terminal will be smaller, as said in the past sections, but the fuel consumption, crew members and O&M of the larger fleet will increase accordingly, compensating the gain.
In Appendix II can be found the same graph, without the cracking additional costs.

4 Transport of Hydrogen by pipelines

4.1 Introduction

The idea of transport hydrogen via pipeline seems fascinating when a large amount of gas has to be delivered, also because of the gained experience on the transport of natural gas along the last decades. Indeed, there is a lot in common among these two technologies, and it would need just some new safety and technical requirements to switch from a natural gas system to a hydrogen system.

Nevertheless, the transport of hydrogen by means of pipeline is not a totally brand-new technology as it could be thought, since it is currently in use worldwide and it finds its routes several decades ago; the first hydrogen pipeline is thought to have been constructed in the 1930's in Germany, and today there are more than 4,500 km of hydrogen pipelines in operation [44]. Just in the United States, there are currently more than 1600km of hydrogen pipelines already in operation; most of them are a direct connection between a hydrogen producer, who own the pipeline itself, and a big hydrogen user.

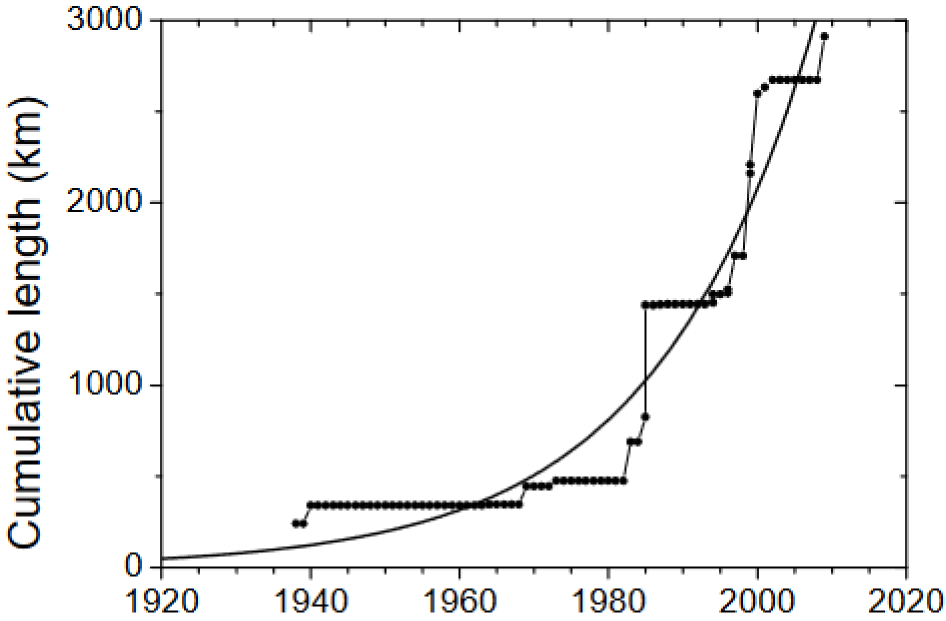


Figure 47 - Global cumulative length of H2 pipelines constructed since 1938 - From "The Cost of Pipelining Climate Change Mitigation" – Graph from 2010

To give an example, Air Liquide has been using a pipeline network in France since the early 1980s to transport high purity hydrogen, at about 100 bar, without showing any operational difficulties and still Air Liquide has a hydrogen pipeline network also in the Ruhr area, which is in operation since the 1930, without reporting any failure or damage too [45].

As can be seen in Figure 47, the currently global cumulative length of the hydrogen´s pipelines globally are still a small sample of what the international transmission market will need in the future. Nevertheless, this small system already proves the feasibility of such technology and it can be though about scaling it up towards international markets´ sizes.

When talking about hydrogen pipelines some of the operative conditions that the literature suggest are: a pressure span within a pipe that generally lies between 2 and 10MPa, the diameter in the range of 0.3-1.5m, and as pipeline materials mainly X42, X52, X56, X60, X65, X70, X80 or other low strength pipeline steels. The expected service life is instead between 20 and 40 years, consistent with the natural gas network.

4.2 Hydrogen applications' safety related issues

The main effects of gaseous hydrogen on pipeline steels are primarily to reduce ductility, reduce fracture toughness, induce a loss of elongation to failure and increase fatigue crack growth rate compared to air or natural gas [46]. Any hydrogen damage requires atomic form to be effective, so it could be assumed that no damage would occur since hydrogen is transported in a molecular way inside a pipeline, but as quantified by the Sieverts´ Law, the molecules of hydrogen can dissociate and then be absorbed into the metallic pipe wall matrix [44]. This phenomenon can lead to a weakening of the steel and an increase in the speed of the propagation of defects as already said. This behavior is generally known as "hydrogen embrittlement".

According to this and based on the literature, when talking about degradation mechanisms of H2 pipelines that can lead to premature failure, the two mainly cited are the heat affected zone degradation (HAZ) and fatigue crack propagation (FCP) in the pipeline base material. The HAZ mechanism can be controlled by the use of X70 or similar steels, which will be treated in the material section, while the FCP appears to be more sensitive [47]. Indeed, once a crack appears its propagation occurs regardless of the type of pipeline steel used, and its evolution is driven by the pressure cycling of the pipe [44]. But different steels do not all have the same sensitivity to hydrogen damages. This sensitivity depends on [48]:

- The diameter,
- The year and the method of the tube's manufacture,
- The purity of the steels in which sulphur/phosphor compounds are present,
- Features arising from welding procedures,
- The purity of hydrogen,
- Operating conditions (mainly high pressure)

Therefore, additional attention should be paid when building a hydrogen pipeline compared to a natural gas one. Indeed, despite high strength pipes are spreading in the NG field, according to [49] there are still some doubts regarding the proper grade of steel, if low or high strength. This because the higher the hydrogen pressure and the higher the strength of the material, the more hydrogen embrittlement and hydrogen-induced cracking will be [49].

4.3 New Hydrogen Pipelines

The construction of new pipelines is unavoidable when the production and the consumption sites are not connected by any existing pipelines, so the possibility of retrofitting or blending is not feasible.

As it happens for the natural gas network, the investment costs involved in the construction of a H2 pipelines depend mainly on four main categories: the material (both quality and quantity), the labor for construction, the miscellaneous and the right-of-way. All of those listed before are features of the system that highly change depending on the specific location of the project, because every project has its own specific set of engineering challenges and specification. This means that the costs are subject to several fluctuation, depending on where the material are bought, the workers are paid, and the pipelines lie. Costs such right of way are extremely dependent on the place where the project takes place and can constitute up to 25% of the total cost of the pipeline, depending on the diameter. For example, a pipeline through a rural area without special environmental concerns can cost five times less than a pipeline of the same length and diameter through a dense urban area [50].

Depending on the diameter and the thickness of the pipe, the relative weight of the 4 categories previously listed could vary. According to [50], the material and labor account for between 65 and 75% of the investment, with the labor having the higher weight of it. Labor and material are indeed the most predictable cost between all of them, since are less sensitivity to the location.

4.3.1 Pipelines Materials

As previously mentioned in section 4.24.2 above, hydrogen has several consequences on the steel properties and according to the operative conditions, the choice of the pipeline´s material could vary. Different standards and guidelines can be followed when choosing the right metal for the pipeline. The most used globally are the ASME B31.12, the API (American Petroleum Institute) SPEC 5 L and the AIGA/EIGA. Generally, they tend to coincide for most of the applications, but they can undergo different tests; indeed, impact testing requirements differ between ASME B31.12 and API 5L, and it is therefore possible that pipelines which were manufactured strictly in accordance with API 5L, do not meet the requirements of ASME B31.12, even though ASME B31.12 includes all the steel pipes in API SPEC 5L. In Figure 48 are shown some properties of the steel pipes that are commonly used in hydrogen applications.

API 5L	X42	X52	X56	X60	X65	X70	X80
Yield strength /Mpa	289.6	358.5	386.1	413.7	488.2	482.7	551.6
Tensile Strength /Mpa	413.7	455.1	489.5	517.1	530.9	565.4	620.6
Allowable Pressure, Max	20.68	20.68	20.68	20.68	10.34	10.34	10.34

Figure 48 - Steel´s pipelines properties according to API 5L

The two-digit number following the "X" indicates the Minimum Yield Strength (in 000's psi) of pipe produced to this grade; this means that, for example, the steel pipe grade X42 has a minimum yield strength of 42 kilo pound/square inch (or 290 Mpa) and X52 has instead 52,000 psi.

In the last years, within the natural gas network, it has been seen a trend where high strength pipes are spreading; this because they allow to use less material for the same safety factor if compared to low strength material, reducing the capital cost even though the material costs is higher [45].

Nevertheless, some literature ([49], [44]) suggest that X42 and X52 type, the low steel grade steel, among the others grades, could be a better fit, because the higher the hydrogen pressure and the higher the strength of the material, the more hydrogen embrittlement and hydrogen-induced cracking occurs [49]. But this point is still controversial, and indeed ASME B31.12 explicitly also allows the use of grades up to X80 for hydrogen service, but the allowable stresses and operating pressures are restricted to such an extent that the cases where X80 is allowed are very limited [44].

In [45], are also mentioned austenitic steels as suitable materials for the pipelines; they have long-term stability in the operating range from 30 to 100 bar and are resistant to embrittlement, but they are much more expensive than conventional steels due to the alloy surcharge, especially due to nickel. In [45], 3600 €/t is set as upper limit for austenitic steel, while 1200 €/t is the cost expected for coated or inliner pipeline which follow the API 5L guidelines. On this regard, [51] proves that the pipelines properties of steel X70 and austenitic stainless type 304L, 316L steels do not undergo any remarkable difference under a hydrogen working environment; so, in the next calculations austenitic steels will not be considered because the higher expenses do not seem to be well motivated. The pipelines´ material is not only function of the diameter, but also of the thickness of the pipeline, which is dependent from the pressure that needs to be hold; the higher the pressure, the bigger the thickness and then the overall cost.

4.4 Hydrogen Pipeline costs

4.4.1 Materials

Regarding the cost of the pipelines, it clearly depends on the quantity of material used; this quantity is function of both the diameter and thickness of the pipe, which rely on the specific use. In much of the literature that can be found online the price of the hydrogen pipelines is often related to the cost function of natural gas pipelines multiplied by a coefficient that accounts for all the differences that the use of Hydrogen instead of natural gas inside a pipeline would bring. This factor may vary from study to study, depending on the assumptions; for example, [50] considers a coefficient of 50% between the material´s cost required by a hydrogen pipeline compared to natural gas, and [26] just the 10%. Nevertheless, [26] is still more expensive because its cost function tend to increase quicker for bigger diameters compare to [50].

Two cost´s functions, based on [50] (violet line) and [26] (blue line), are shown below in Figure 49:

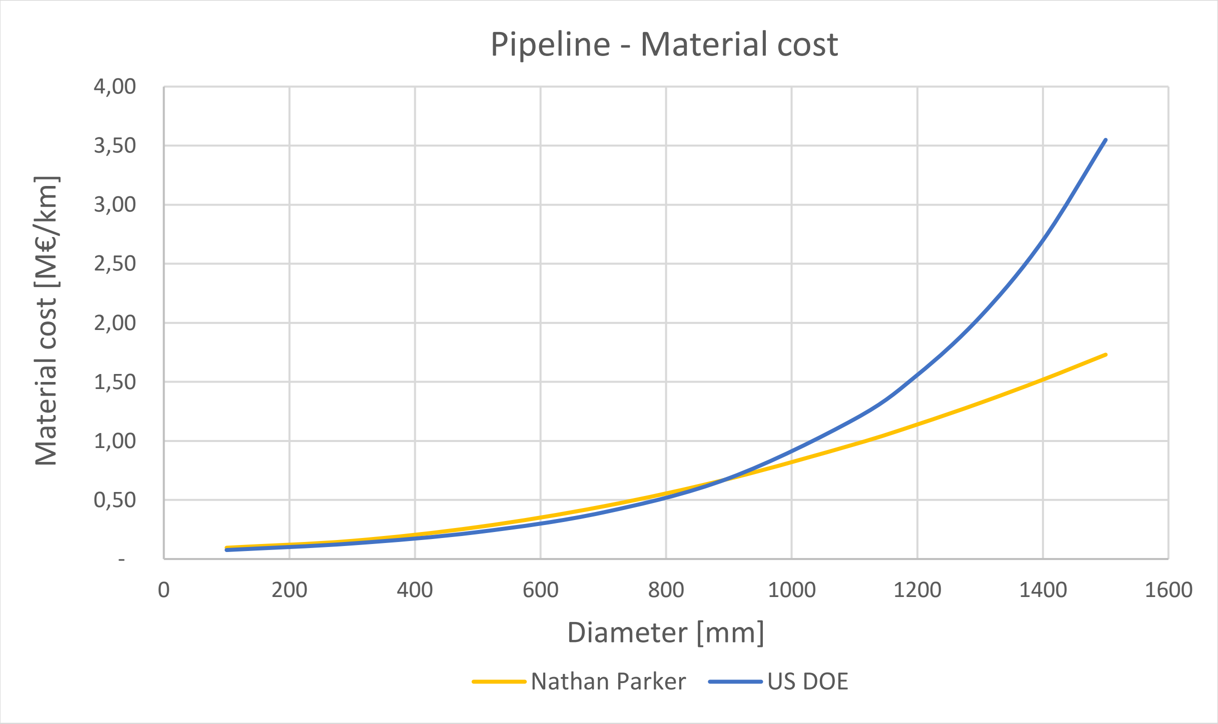


Figure 49 – Pipelines´ material cost

For the next calculations, the US DOE blue line from [26] will be chosen, since it is the most recent one between them and it has a more reasonable behavior for larger diameters, since the relationship between a pipeline´s diameter and its cost should stay quadratic within all the diameter´s range choice.

The type of steel used in these studies is not known, but based on what [11] states, the costs´ difference between steel X52 and X72 is just about 10%, so it is considered to have little influence on the final investment; therefore, the research of the type of steel is not further investigated.

4.4.2 Labor

The labor cost consistently averages between 40 and 50% while rest (material, miscellaneous) vary greatly depending on diameter. It consists mainly of the work required to put in place the pipelines and dig the ground for it; therefore, it involves all the workers, machineries and fuels spent in such a job.

Even for the labor, [50] (violet line) and [26] (blue line) give a cost function, shown as follows in Figure 50:

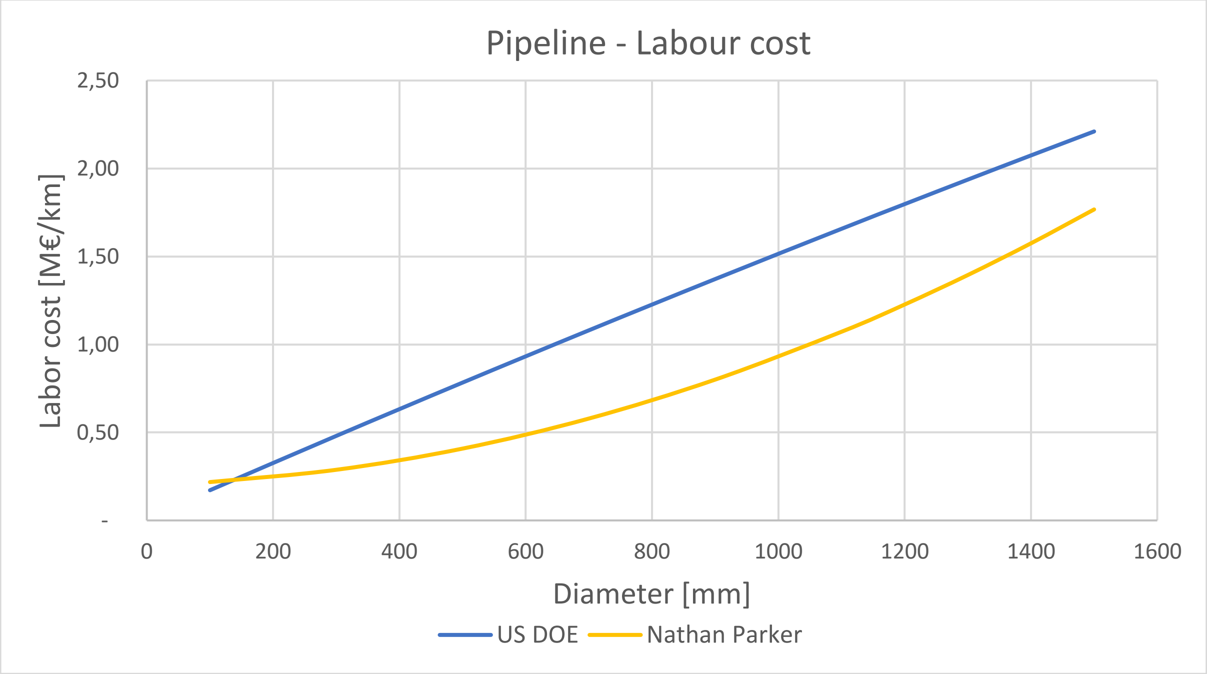


Figure 50 – Labor cost functions

The blue cost function from [26] is again chosen for the following calculations, because of the same reasons of the material´s cost choice.

4.4.3 Right-of-way

The right of way cost depends on the pipe diameter and is a small fraction of the transmission pipeline construction´s cost. Since the right of way costs are very hard to predict by just the length of the pipeline because they are too much sensible to the location chosen, they will not be considered.

4.4.4 Miscellaneous

Miscellaneous costs are all costs not included in labor, material, or right of way. They generally include surveying, engineering, supervision, contingencies, allowances, overhead, and filing fees [50]. The miscellaneous, or administrative and engineering costs, can account for up to 30 percent of the total construction cost. Generally, these costs depend on the size of the project, diameter and length of pipeline [50]. Looking at Figure 51, the possible cost functions based on [50] (violet line) and [26] (blue line) are plotted.

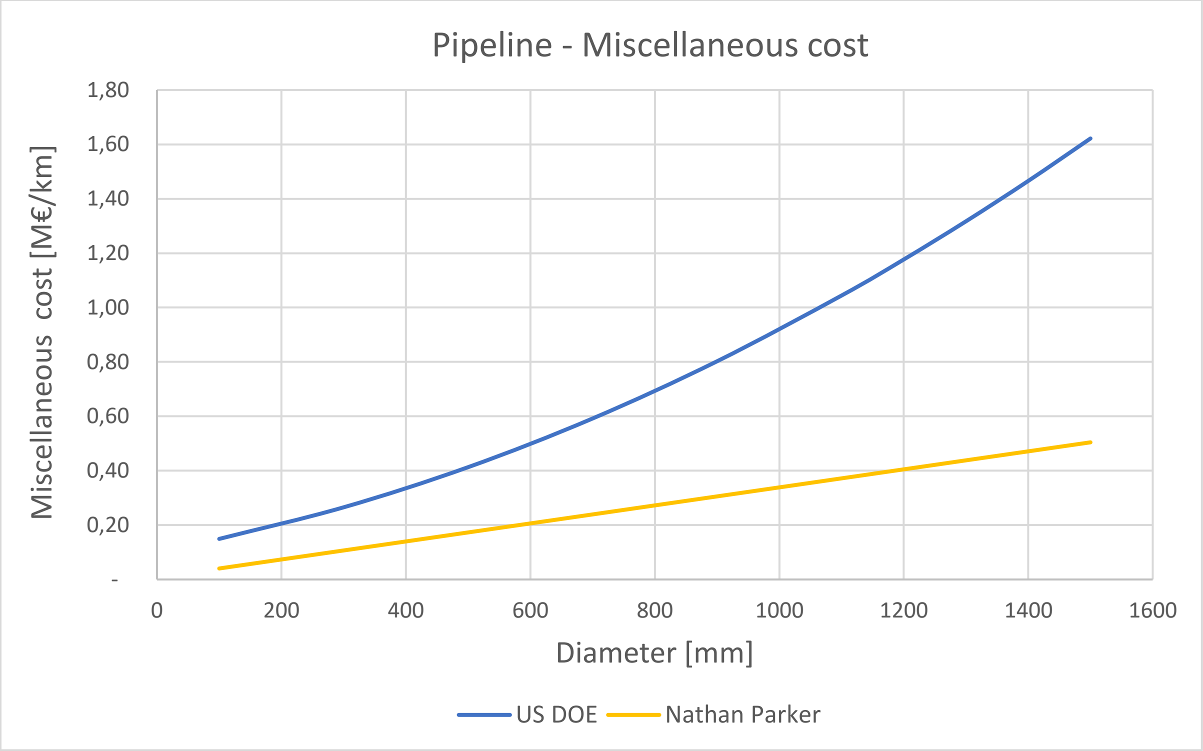


Figure 51 - Miscellaneous costs

For the same reasons of labor and material´s cost, the blue line from [26] is going to be used for the calculations.

4.5 Dimensioning of the pipeline diameter

The diameter of a gas network has the main impact on the capital expenditure of the system, so it is a key point to size it properly to both minimize and optimize the overall costs. To evaluate it, equation (4.1) from [52] it has been used, which links all the key parameters and allow to estimate the diameter given their values. Equation (4.1) can be then manipulated, to make the variable D explicit instead of Q.

$$Q = C.D^{2.5}.e.\sqrt{\frac{(p_1^2 - p_2^2)}{d.Z.T.L.f}}, \tag{4.1}$$

Where:

- C: proportional constant, dimensionless. Equal to 0,000129
- Q: volumetric flowrate, [Nm3/h]
- e: pipeline efficiency, between 0,9 and 0,95
- P_1: beginning of pipe pressure [bar]
- P_2: end of pipe pressure [bar]
- d: hydrogen relative density to air, dimensionless
- Z: Average compressibility factor [dimensionless]: Typically, Z = 1.0 - 1.4 for H2 in the pressure and temperature ranges examined
- T: Gas temperature [K]
- L: length of pipe [km]
- f: friction factor, dimensionless.

Given a fixed distance L, usually the one between two compression stations, the choice of the diameter is mainly function of the pressure drop inside the pipeline and the flowrate within that distance, other than the properties of the gas itself. Once we have the diameter function it is easy to plot its behavior; here is shown in Figure 52, assuming a fixed distance between compression station of 250km:

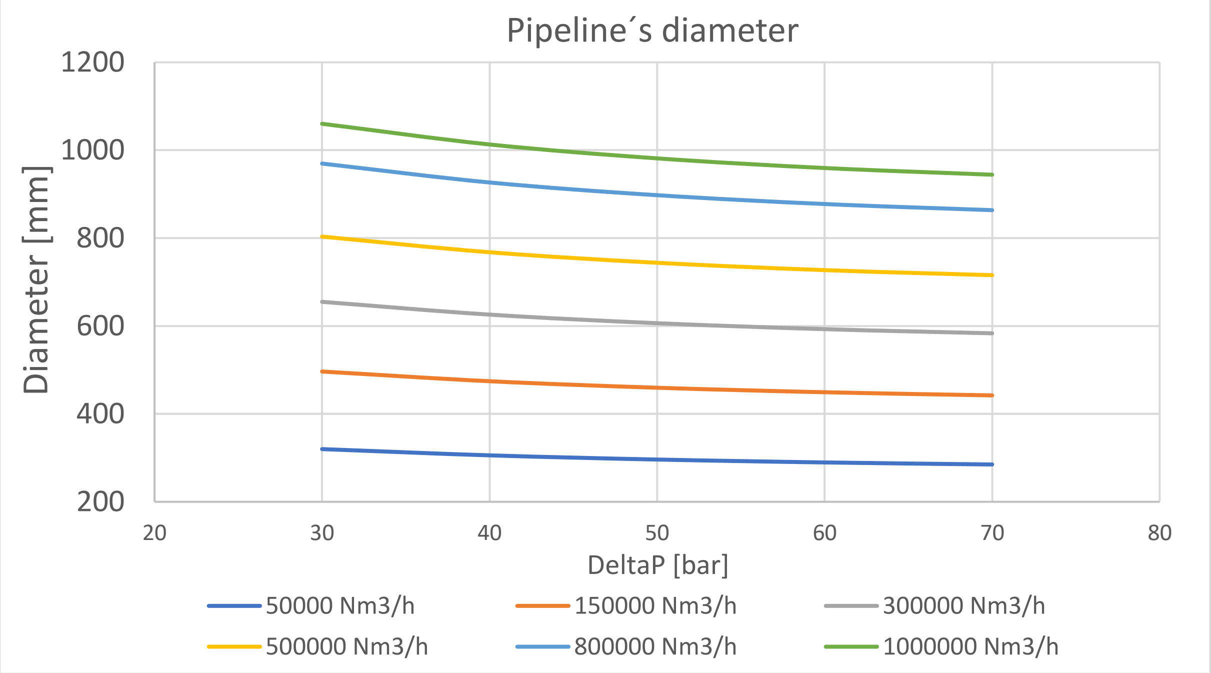


Figure 52- Changing of the diameter´s size depending on flowrates and pressure drop

The main outcomes from Figure 52 are:

- Given a volumetric flowrate, lower diameters imply higher pressure drop inside the pipeline. This means that some money can be saved for the purchase of the pipeline´s material, but higher expenses for the compressor should be considered, due to the bigger size of the compressor and the higher energy required
- Fixed the pressure drop, the higher volumetric flowrate the bigger the diameter

4.6 Compression station

Generally, a transport network mainly consists of pipelines, compression stations and pressure-reduction stations. In this section the focus will be on the compression stations.
Compression stations can be approximated as a system with two main features: the compressor and the equipment that gives the compressor the required energy, which ~~in turn~~ can be either a gas turbine or an electric motor.

In nowadays gas pipelines, the most common choice is to use gas turbines to be independent from possible network instabilities (ex. blackouts), which would lead to the stop of the compression work and transmission operation; on the other hand, in some recent literature, is taking place the idea to substitute the turbines, at least at the first stage, with electric motors because it is considered as more reliable and efficient than either reciprocating engines or gas turbines with a faster ramp up [53]. Beside this, the research on hydrogen turbines is still going on, and it will need some years still to be ready for the market. From a strictly cost point of view, an electric motor would cost much less than a gas turbine, \$30 to \$50/kWe versus \$400 to \$600/kWe according to [54], but the availability and reliability of grid electricity is still the biggest concern when using electrically driven compressor [53]. When a gas turbine is the compressor driver it has to be noted that, if technical reasons lead to use reciprocating compressors, use of a large step-down gear box is mandatory, because reciprocating compressors needs low driving velocities to work properly, unlike centrifugal compressors [4].
Compressors in hydrogen´s pipelines, ~~t~~are one of the main network components, both from a technical and economic point of view, so it is worth to spend few pages on current criticisms and possibilities regarding this topic.

The state of the art of hydrogen compressors is not ready yet for the future market needs, but it is in constant evolution. Nowadays, hydrogen compressors may be available, but are still one of the main sources of failure in hydrogen stations [47].

Given the actual pressure ratio and flowrate needed in the system, the state-of-the-art of compressors for transmission´ systems, are multistage oil-free reciprocating machines, but additional research and development activities are under way in both the private and public sectors. Generally, such a technology has high maintenance costs due to wearing components such as valves, rider bands, and piston rings [55].

Hydrogen compressors are also expensive due to the required materials, the physical size needed to supply the needed compression power, and the redundancy to provide reliability. Hydrogen compressors must also have tight tolerances and special seals to withstand high pressures [55].

Compressing hydrogen, compared to other gases and especially natural gas, faces several technical challenges: considering the low molar weight of hydrogen, its recompression ratio is four times lower than for natural gas for a given rotor speed, implying a larger number of stages. Moreover, hydrogen compressors usually cannot achieve higher pressure ratio than 1.1-1.2 per stage [56], [57]. On the other hand, fixed the pressure ratio, the rotational velocity to compress hydrogen should be between 1.74 [55] and 3 [58] times higher than natural gas compression, reaching velocities larger than 700 m/s [58]; This implies the research of new hydrogen-resistant impeller materials able to withstand high centrifugal forces since high velocities could cause problems to the impeller of the compressor itself [55].

Depending on the specifications of the system, pressure ratios and flowrate above all, different compressor´s choices can be taken. But to fulfill the requirement for rugged and reliable operations with large volume flows, as a pipeline´s system needs, the selection is going to fall only on either reciprocating pistons or centrifugal compressors for most of pipeline transport and storage application [56], as can be seen in Figure 53:

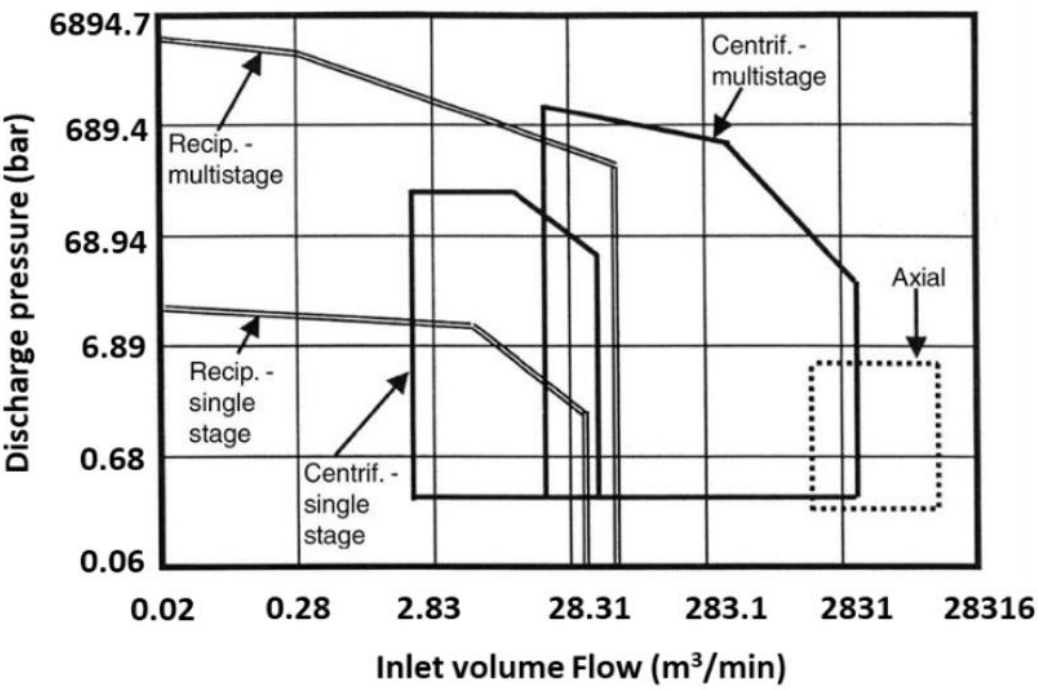


Figure 53 - Hydrogen´s compressors main decision´s leading characteristics

4.6.1 Reciprocating compressor

Reciprocating compressors are ideal for low to moderate flow and high-pressure application as showed in Figure 53. They are typically used in refineries and chemical plants to compress industrial grade H2 to high pressure. Nonetheless, oil free H2 reciprocating compressors are not exempt from embrittlement and so they experience frequent failure of the sealing rings due of non-uniform pressure distribution. Furthermore, thermal protection of various components becomes critical in this kind compressors because there is no oil to act as heat sink [53]

Reciprocating compressors possess a higher number of moving parts with respect to centrifugal compressors, produce more vibrations and require more maintenance [59].

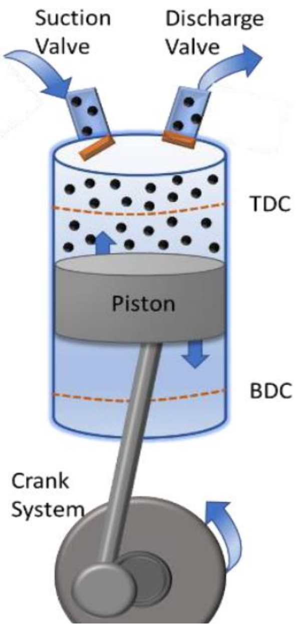


Figure 54 - Scheme of a reciprocating compressor

4.6.2 Centrifugal compressor

Regarding centrifugal compressors there are some discrepancies on the available literature. For example, [59] states that centrifugal compression is impractical for H2 because of its small molecular size, since compressing and pumping H2 requires a different approach from that of heavier gases and [55] claims that centrifugal compressors create more operating problems than reciprocating compressors. [56] instead says that "the efficiency of a compressor is determined by a basic aerodynamic blade design and can be optimized for any type of gas. Thus, "there is no reason a centrifugal compressor cannot be designed to operate efficiently for a light gas." Also [53] agrees with [56] pointing centrifugal compressor out as feasible alternative for high hydrogen flowrates. Indeed, according to [56] centrifugal compressors are preferable for hydrogen compression because of reduced environmental leakage, no piping pulsation and vibrations, and overall lower maintenance costs, even compared to reciprocating compressors.

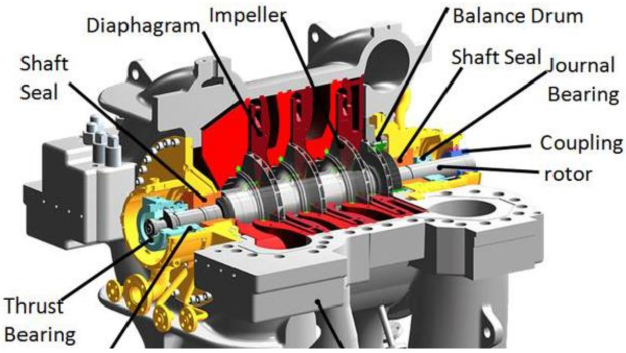


Figure 55 - Scheme of a centrifugal compressors

Centrifugal compressors are dynamic compressors that are most commonly used in applications that require high throughput and moderate compression ratio. They compress the process gas using a rotating impeller with radial blades that imparts kinetic energy to the process gas by increasing its velocity. By means of a diffuser, the kinetic energy is then converted into pressure. Since the compression ratio largely depends on the molecular weight of the gas, centrifugal compressors will require impeller tip-speeds around three times higher for H2 than those used for natural gas. [53]

Summarizing, for high discharge pressure and medium mass flowrates, the best option would be a reciprocating compressor, while for high mass flowrate and medium compressor´s outlet pressure the centrifugal compressors seem to be the recommended choice. In a gas transmission system where the pipeline diameter can lie between 0,3 and 1,5 meters, the mass flowrates can reach several thousands of hydrogens cubic meter per minute; therefore, in this analysis the compressors will be always assumed as centrifugal. Anyway, for both technologies, the life expectancy is estimated between 10 and 15 years.

The annual operative expenditures of a compression station account for the purchase of the energy required by the transport of hydrogen. The energy needed to run the compressor could be provided either by an electric motor or by the mechanical work that comes from a gas turbine.
As it has been said in the last section, in our assumption the compressor driving force will be an electric motor, so all the calculations are going to be performed accordingly.

There is a very spread agreement ([60], [61], [47], [45]) where the energy consumption range of a compressor should lie: between 1.3%, as a minimum estimation, and 3%, as a maximum, of the energy content of hydrogen (33,33 kWh/kg). This percentage can be then translated in kWh/kg, being 1,3% equal to 0,43 kWh/kg and 3% to 1 kWh/kg. To do a comparison with natural gas, the energy required is roughly around the 0.3% of the energy content of the compressed methane, so the hydrogen compression operation is more energy intensive than natural gas.

Equation (4.2) shows the specific energy consumption of a compressor, by means of a polytropic and adiabatic function:

$$Wc = \frac{N * Z_{average}}{3600} * \frac{1}{\eta_{is} * \eta_{mec}} * \frac{\gamma}{\gamma - 1} * r * T_1 * \left[\left(\frac{Pc}{Pi}\right)^{\frac{\gamma-1}{N+\gamma}} - 1\right] \tag{4.2}$$

Where:

- N: number of stages
- γ: Ratio of specific heat under constant pressure (Cp) to specific heat under constant volume (Cv): $k = Cp/Cv = 1.41$
- Z: Average compressibility factor [dimensionless]: Typically, Z = 1.0 - 1.4 for H2 in the pressure and temperature ranges examined
- T_1: Suction/inlet temperature (K);
- P_i: Suction or inlet pressure (bar);
- P_c: Discharge or outlet pressure (bar)
- r: Universal gas constant, R = 8.314 J/mol.K
- η_{is}: isentropic efficiency
- η_{mec}: mechanical efficiency

Based on equation (4.2) the specific energy consumption of a compressor is mainly dependent on the pressure drop that the compressor has to provide to the network (P_c/P_i), on the properties of hydrogen (Z, γ) and on the number of stages.

Plotting this equation with regard to different pressure drops that the compressor has to overcome, it fits well the range (1,3-3%) cited before just for low suction pressure, as can be seen in Appendix III - Figure 85. Nevertheless, when the inlet pressure of the compressor work is increase, the compressor work reaches values too low, which lead to too optimistic consumption values (0,04 kWh/kg for a pressure drop of 10 bars). To overcome this problem, it has been decided to estimate and then use for the calculation, an equation that lies exactly on the pressure range already discussed and which varies accordingly to the pressure drop demanded to the compressor, Figure 56.

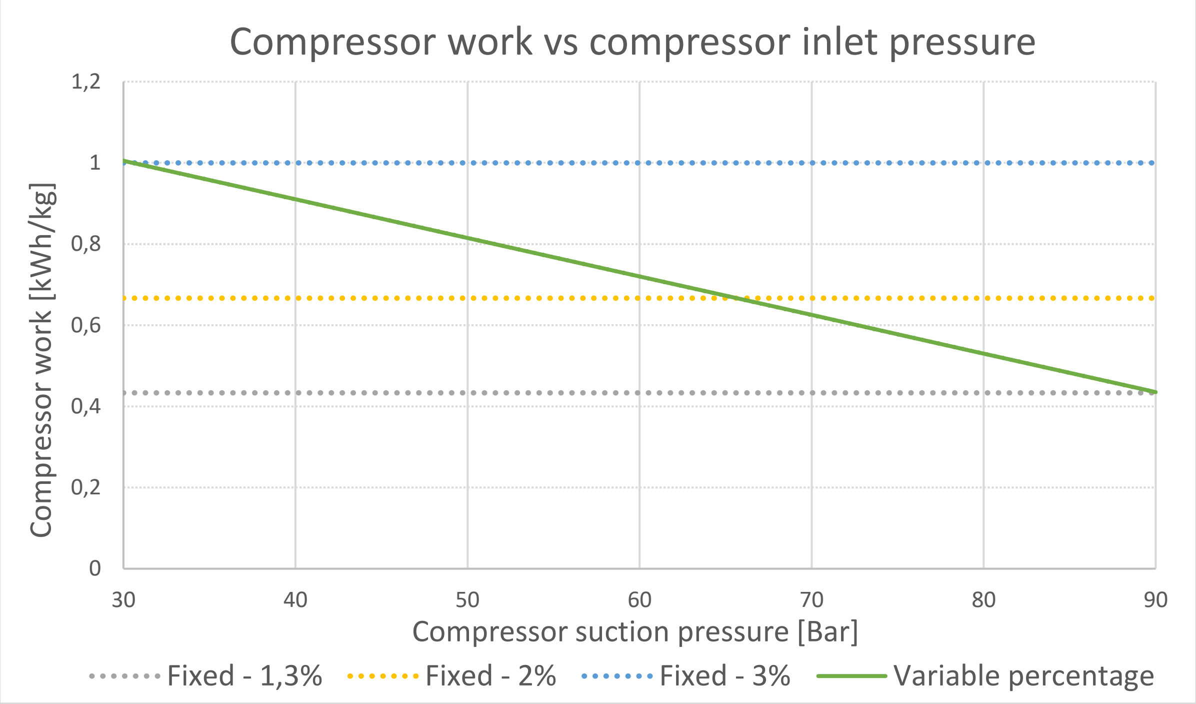


Figure 56 – Energy required by the compressor

In this case the outlet or discharge pressure is fixed to 100 bar, so the pressure drop is 100 – X, where X is the suction pressure plotted on the X axis.

The mathematic estimation of the compression OPEX it is based on the following formula (4.3):

$$OPEX_{comp} = \frac{CostOfElectricity * W_{compressor} * m_{annual}}{\eta_{elec.motor}} \tag{4.3}$$

Once the annual mass “m_{annual}" ($[kg]$) and the "$W_{compressor}$" are known, to finally get an estimation of the annual operative expenditures, we still need to make some hypothesis about the electric motor efficiency and the cost of the electricity, which are listed in Table 12.

Table 12 - Compressor OPEX hypotheses

Hypothesis	Symbol	Value	Unit
Electric motor efficiency	$\eta_{elec.motor}$	98	%
Cost of electricity	$CostOfElectricity$	60	€/MWh

The capital expenditure for a compression station are mainly attributed to the purchase of the compressor and its driver, either electric motor or gas turbine. Both, the compressor and the driver, are function of the power required and their prices can start from few thousand up to millions of euros.

Compared to natural gas networks, considerably higher compression power is needed, because compressors operate on the basis of volume rather than energy content [55].

On the available literature ([62], [53], [63]) different equations are suggested to get an estimation of the compressor power required, based on the pressure drop that the compressor has to provide to the network (P_c/P_i), on the volumetric flowrate (Q), on the properties of hydrogen (Z, γ) and on the number of stages. All of them can be found in Appendix III, equation (3.2), (3.3), (3.4) and displayed in the same Appendix in Figure 86. To be conservative, in the model it has been taken the highest value ([63]) among the three found for the compressor power, and it is plotted below in Figure 57 for different volumetric flowrates.

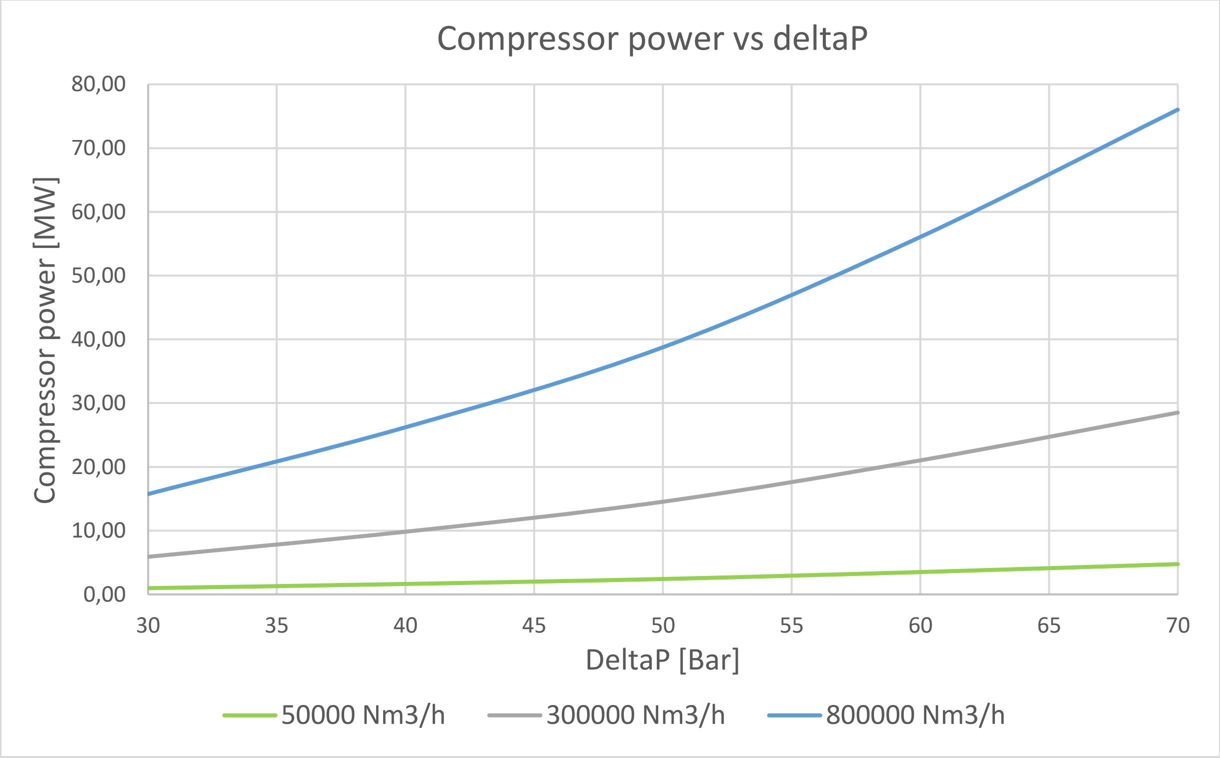


Figure 57 - Compressor power based on [63]

As showed in Figure 57, the volumetric flowrate has a huge weight on the final power required by the compressor, and it is therefore one of the main parameters for its sizing.
Academic literature and industry both use empirical cost correlations to determine how much a compressor will cost based on its size in KW or MW [53], so once the power required is known, the capital expenditure can be deducted.

Even in this case, the literature ([64], [26], [65], [66]) provides different cost functions and values which are simply consistent among them, with just few discrepancies; they can be seen in Appendix III, Figure 87.

It has been picked the most recent reference [65], which is also the cheapest one, and is shown below in Figure 58:

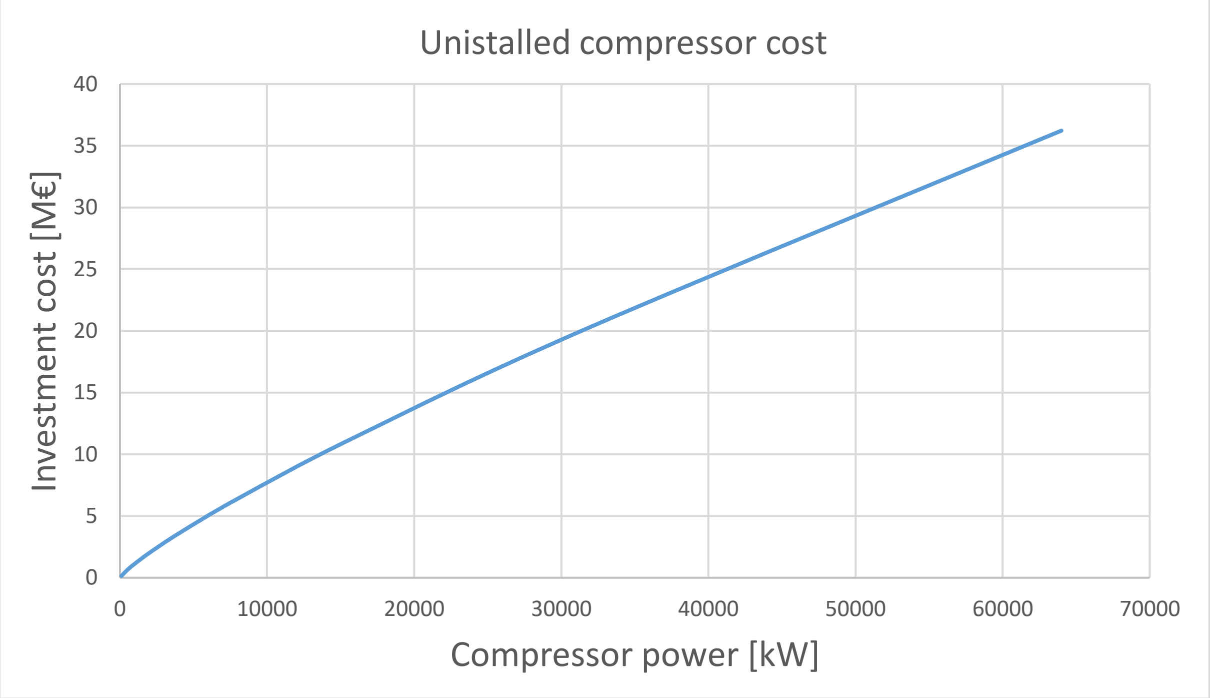


Figure 58 - Uninstalled compressor cost based on [31] cost estimation

It is worth to mention that these costs are for uninstalled compressors; according to most of the literature, these values must be multiplied by a factor of 2 to get the desired installed investment cost of the compressor.

Based on the trend seen for the compressor´s power (Figure 86) and the compressor´s cost (Figure 58) it can be state that the relation between compressor cost and pressure drop is nonlinear; implying that a big pressure gap leads to much higher fixed costs [45].

4.7 Overall Pipeline capital expenditures

We have now seen all the components that are included within the estimation of the pipelines´ capital expenditures. While the total cost of the compressors´ purchase depends on several factors, like its size, the total length of the pipeline and the distance between two compression stations, all the other costs (material, labor, miscellaneous) are mainly dependent on the pipeline´s diameter; that is why in Figure 61 the aggregated costs of these 3 components, with respect to the diameter, are displayed:

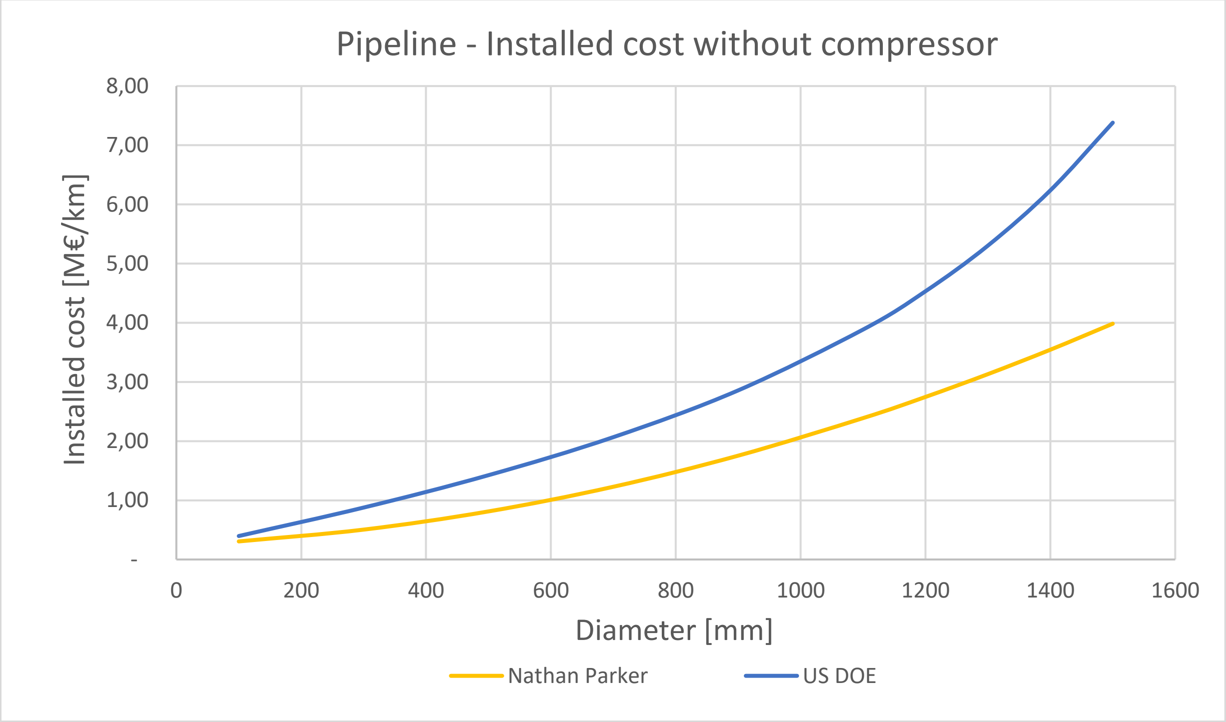


Figure 59 - Aggregated pipeline´s capex without compressor costs

The blue line [26] is the cost function that will be adopted for the calculations in the next sections. Looking at Figure 60 - Pipeline investment breakthrough Figure 60, it is possible to get an idea of how material, labor and miscellaneous weight on the final cost of a pipeline. This is just an example made for a specific case (D equal to 590mm), since these values change depending on the diameter´s size.

In Figure 60 is not shown the right-of-way because is the most unpredictable factor, which can vary significantly depending on the site of the project but can reach even the 20% of the total investment.

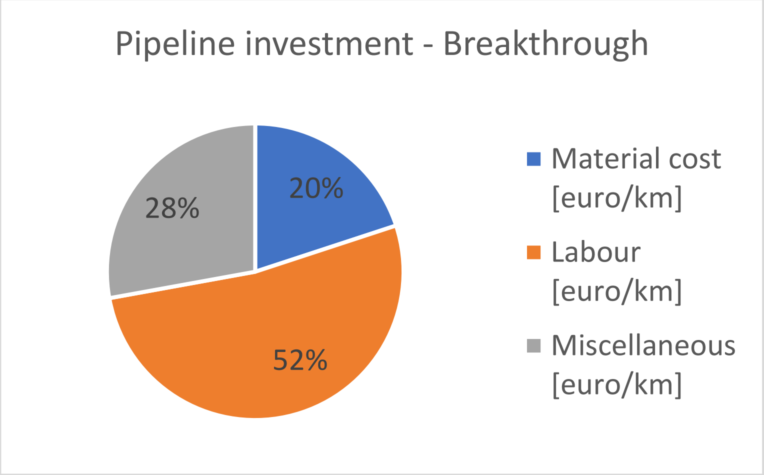


Figure 60 - Pipeline investment breakthrough – Diameter equal to 590 mm

During the bibliography research, it has been found in [45], different cost functions for the capex of the pipeline worth to compare with this analysis results. [45] accounts for material, labor and miscellaneous costs and also the compressors capex, assuming a fixed distance between them of 250 km. Considering the same assumptions is then possible two compare the results of this analysis and the outcome from [45], which are shown below in Figure 61.

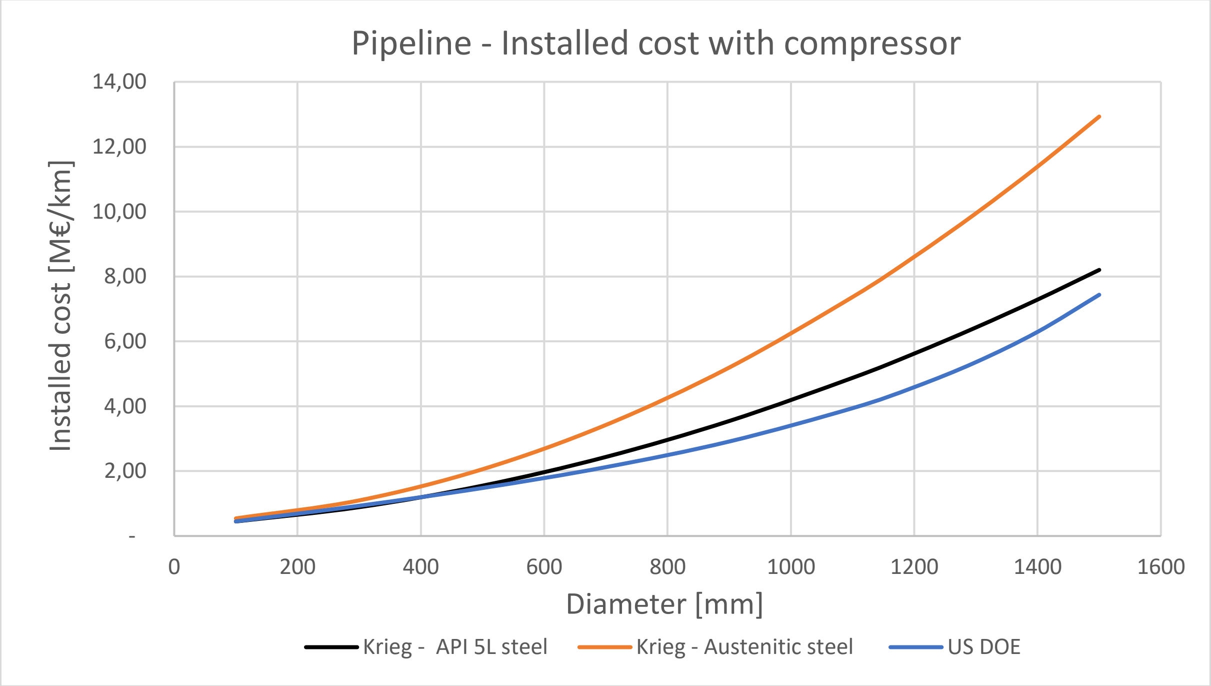


Figure 61 - Aggregated pipeline´s capex with compressor costs

[45] Proposes two cost functions, one cheaper that the other, for two different materials: austenitic (red line) and common steels (black line). As it has been said in the materials´ section, austenitic steels are much expensive with respect to common steels that follow the API 5L guideline and can be seen easily in Figure 61.

Since in this analysis austenitic steels are not considered, the focus must rely on the black and blue lines. Nevertheless, the black line can be useful only when the compressor distance is equal to 250 km. To have more design flexibility on the compressors distance choice, the blue line is going to be considered for economic assessments.

4.8 Pipeline Operative Expenditure and Operation and Maintenance costs

The OPEX of the gas network overlap with the OPEX seen in the compressor section, because the only operative expense is the energy required by the compressor to run, therefore no further analysis will be performed on this regard.

The operation and maintenance (O&M) are instead function of compressors and pipeline capex, and their values are shown in Table 13:

Table 13 - O&M hypotheses

Hypothesis	Value	Unit
Compressor O&M	4% [65]	% of the Capex
Pipeline O&M	1% [61]	% of the Capex

4.9 Levelized Cost Of Hydrogen (LCOH)

The Levelized Cost Of Hydrogen (LCOH) is a measure that accounts for all the project lifetime expenses divided by the goods the project is aimed for (in this case the amount of hydrogen delivered). It is expressed in €/kg and for this reason make possible the comparison among different technologies. More in detail, is evaluated dividing the sum of all the yearly annualized CAPEX, the OPEX and the O&M costs, by the annual amount of hydrogen traded, as it is shown in equation (4.4).

$$LCOH = \frac{\sum_i \frac{CAPEX_i}{a.f._i} + OPEX_i + O\&M_i}{H2\ treated\ annually} \tag{4.4}$$

Where the actualization factor $a.f.$ is equal to:

$$a.f. = \frac{\left(1 - \left(\frac{1}{\left(1 + \frac{r}{100}\right)^{l.t.}}\right)\right)}{\left(\frac{r}{100}\right)} \tag{4. 5}$$

With:

- r = discount rate
- l.t. = lifetime [years]
- H2 treated annually = [kg/year]

In the previous sections all the components of the LCOH (Capex, Opex and O&M) have been investigated, assigning to each of them a specific investment or a cost function. To finally be able to assess which is their weight on the final cost of hydrogen, Figure 62 shows their share on the total LCOH. Clearly, the investment cost of the pipelines is the main cost driver, but a particular attention should be assigned to the compressors: summing the initial investment, the cost of energy and the replacement costs due to their short life, almost the 40% of the total LCOH is reached.

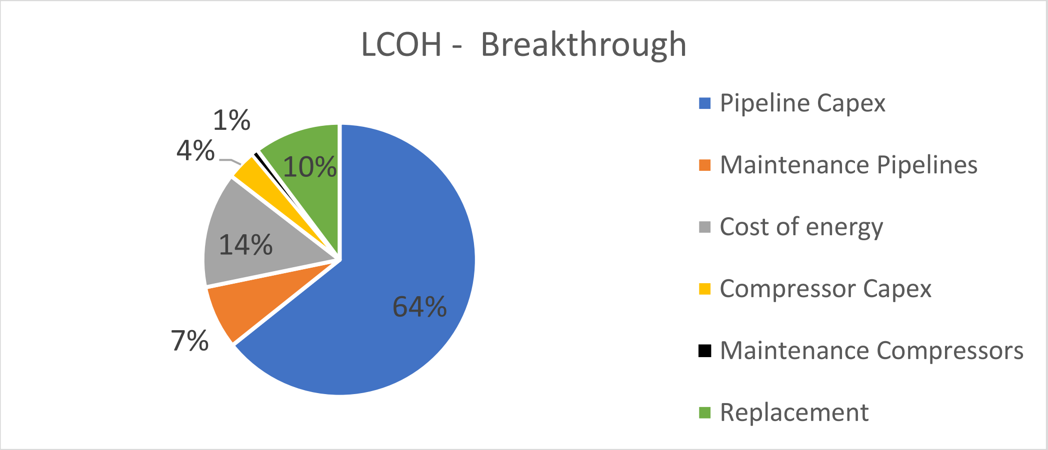


Figure 62 - LCOH Breakthrough

As stated at the beginning of this chapter, it does not exist one project equal to the other when it comes to the construction of the pipelines, because every project has its own specific set of engineering challenges, specification, and hypothesis. Two equal systems, but with different assumptions and input data, could lead to outcomes far from each other. That is why in this part the main hypotheses and input parameters that affect the system are reported in Table 14 to give a simpler way of comparison with other models, if needed.

Table 14 - General hypotheses and input parameters of the pipeline system

Hypothesis / Parameter	Value	Unit
Pipeline lifetime	40	years
Pipeline utilization factor	60	%
Discount rate	8	%
Electrolizer output pressure	30	bar
Pipeline nominal pressure	100	bar
Cost of electricity	0,06	€/Kwh
Compressor pressure drop per stage	1,2	-
Compressor lifetime	10	years

4.9.1 Sensitivity analysis

Pipelines are usually considered as the cheapest alternative when the quantity of hydrogen transported is really high and the transport distance is short. That is why in Figure 63 is plotted the behavior of the LCOH assuming different values for the transport distance and rate of production.

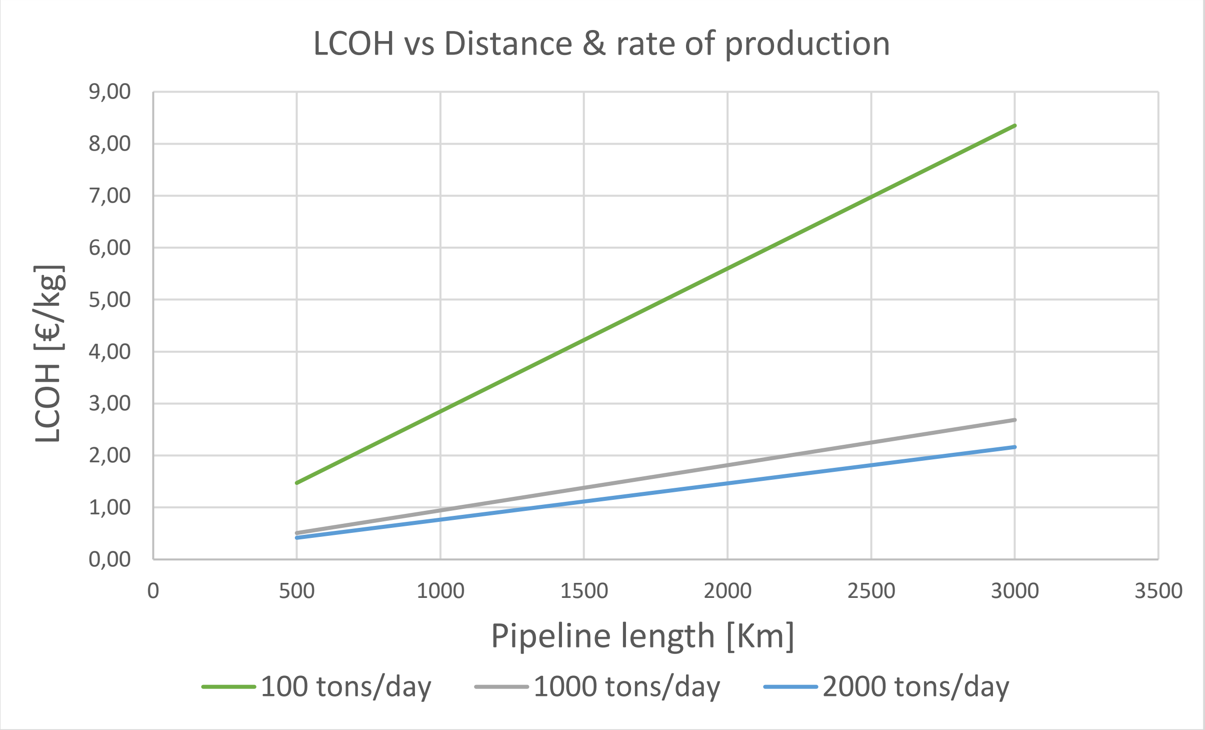


Figure 63 - The value of LCOH varying transport distance and rate of production

The distance has a massive impact on the LCOH, even though this effect is mitigated by the increase of hydrogen production: assuming a production rate equal to just 100 tons per day (Orange line) the LCOH for 3000km of transport is 8 €/kg, more than 5 times compared to 500km of transport (1,58 €/kg); on the other hand, for a production rate of 2000 tons per day (Blue line), the LCOH is 2,11 €/kg, 4 times bigger for 3000km than for 500km (0,5 €/kg). Therefore, it can be stated that the higher the hydrogen production, the less the distance influences the final LCOH.

Another design cost-driver is the spacing between hydrogen compressors along a pipeline. It is likely that the spacing between hydrogen compressors would fall between 100 and 600 km [55], [61], but the choice of this parameter has several and significant implications on the system, which would be easier to explain while looking at Figure 64.

A sensitivity analysis of the LCOH has been performed assuming different distances between compressors, respectively 100 km (Orange line), 250 km (Green line) and 500 km (Violet line), as functions of different pressure drops along the pipeline.

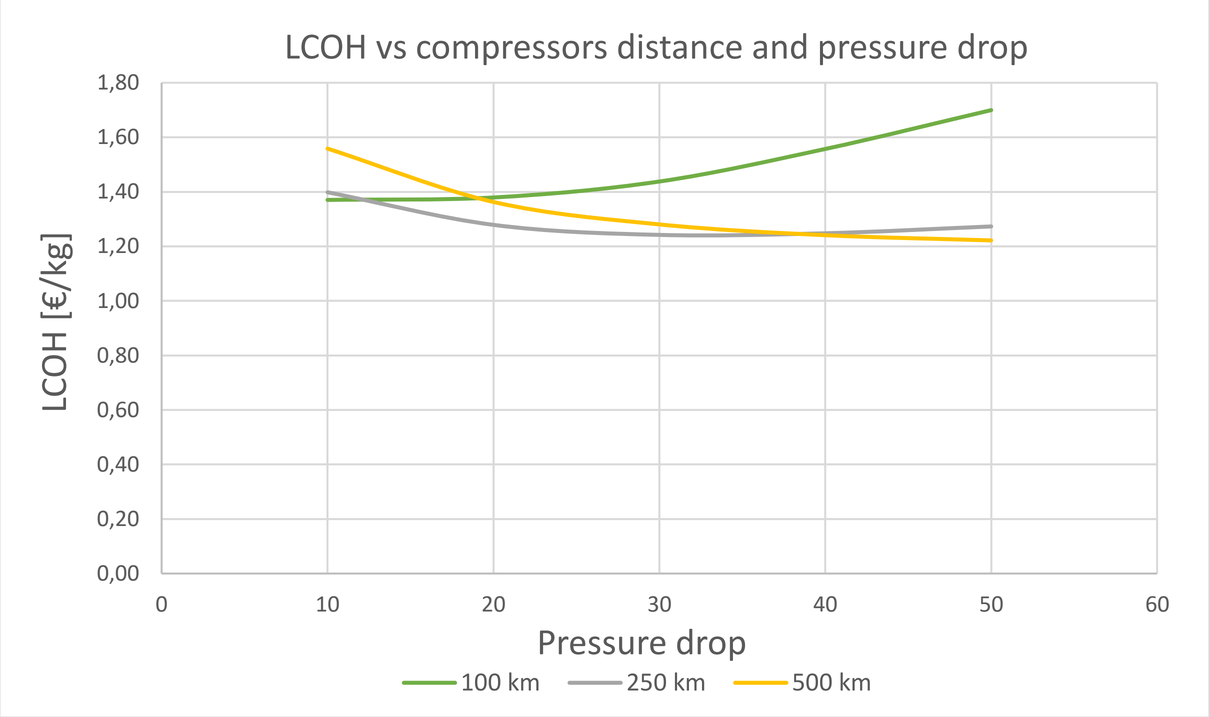


Figure 64 - The LCOH depending on the distance between compressors and different pressure drops

In the example of Figure 64, the production is fixed and equal to 500 tons of H2 per day. Let´s firstly focus on just one line, the grey one for example, which stands for a distance between compressors of

$$Q = C.D^{2.5}.e.\sqrt{\frac{(p_1^2 - p_2^2)}{d.Z.T.L.f}},$$

250km. As it was shown in equation (4.1), that is recalled below, the pressure drop is mainly function of the diameter and the flowrate, because the pipeline and hydrogen properties are usually assumed as constants. Therefore, allowing the pressure drop to increase within the 250 km from 10 bar to 50 bar, is the same as saying that the diameter is getting smaller for a constant flowrate. Small diameters mean lower capital expenditure for the pipeline, but at the same time higher installed power and required energy from the compressor during its operation.

Said so, the LCOH along a line with fixed compressors´ distance is a trade-off among all the components said previously; in the green line´s case, the optimum lies at a value of 30 bar pressure drop, meaning that for such a pressure drop the higher compressor´s capital and operative expenditures are overcome by the savings in the pipeline investment.

Finally, for the specific example shown in Figure 64, the best option would be to put a compression station every 250 km, which has a related LCOH of approximately 1,22 €/kg.

5. Case study

The goal of this section is to provide the best solution as a way of transport hydrogen among all the three technologies faced in the previous chapters (Liquid hydrogen, Ammonia and Pipelines), for a given hypothetic scenario. As a case study, the transport from Casablanca (Morocco) to Marseille (France) is considered. Morocco is of interest because of the future cost of hydrogen production from photovoltaic energy, that is forecasted to fall down to 2,5-3 €/kgH_2 in the near future. On the other hand, Marseille is trying to arise as one of the main hydrogen-hub in Europe, sponsored by the French government. Moreover, this case study makes possible the comparison between the ship transport for LH_2 and NH_3, and the transport of hydrogen by pipeline.

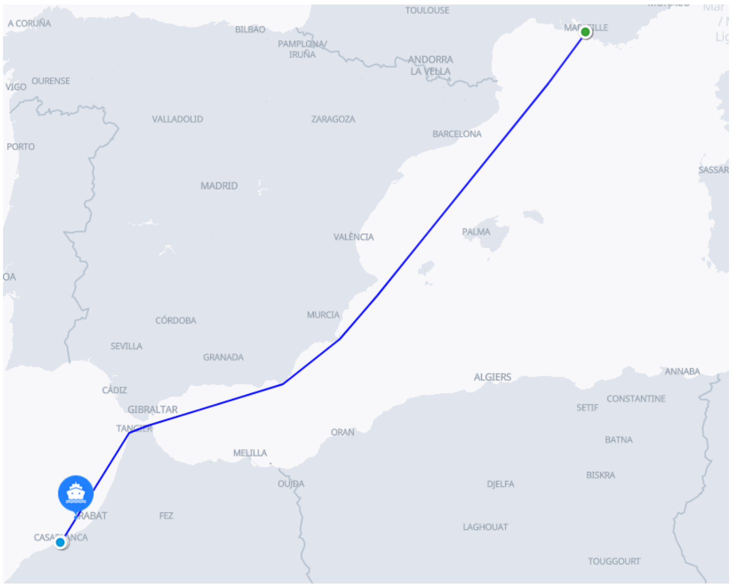


Figure 65 - Possible shipping route between Casablanca and Marseille

In Figure 65, the hypothetic shipping route is shown. For the transport by ships, the distance is estimated as 1640 km between the two ports of Casablanca and Marseille, while for the pipelines the distance is assumed to be larger than the case of shipping transport, because natural boundaries like hills or mountain do not allow a straight path.

The Hydrogen production is the same for all the technologies and fixed to 500 tons of Hydrogen per day.

In Table 15 are reported the main assumptions for the 3 technologies.

Table 15 - Case study hypotheses

Hypothesis	**LH_2**	**NH_3**	**Pipeline**
Hydrogen daily production [tons/day]	500	500	500
Distance [km]	1640	1640	2000*[2]

---

[2] *40 km of pipelines are underwater

5.1 Liquid Hydrogen

For what concern the Liquid Hydrogen case, once the distance and the production are fixed, the only degree of freedom for the best system´s design choice relies on the number of ships involved in the fleet. For this reason, in Figure 66 the evolution of the LCOH depending on the dimension of the fleet is plotted.

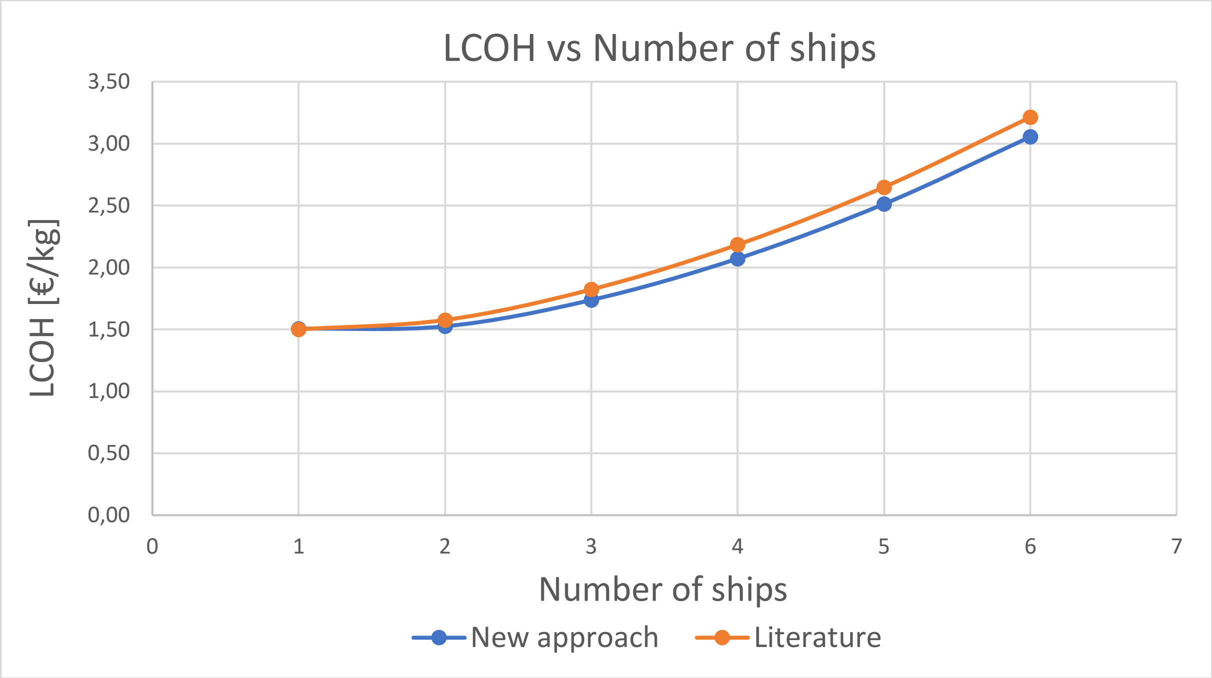


Figure 66 - LH2 LCOH depending on the number of ships

Given this specific situation, the best configuration for the Liquid Hydrogen supply chain would be the purchase of only one ship, according to both the literature (orange line) and new approach (Blue line). Since the overall system is defined, the LCOH for the LH_2 can be evaluated.

In Figure 67 is possible to see how the LCOH is divided in its main components. Over a total cost of 1,52 €/kgH_2, the largest weight belongs to the liquefaction plant, 0,92 €/kgH_2, with both capital and operative expenditures included. The liquefaction plant is then followed by the ship, and the terminals specific costs, respectively 0,21 €/kgH2 and 0,2 €/kgH_2.

This configuration is not general and changing even one between distance of transport and daily production could lead to another choice of ships and final LCOH value.

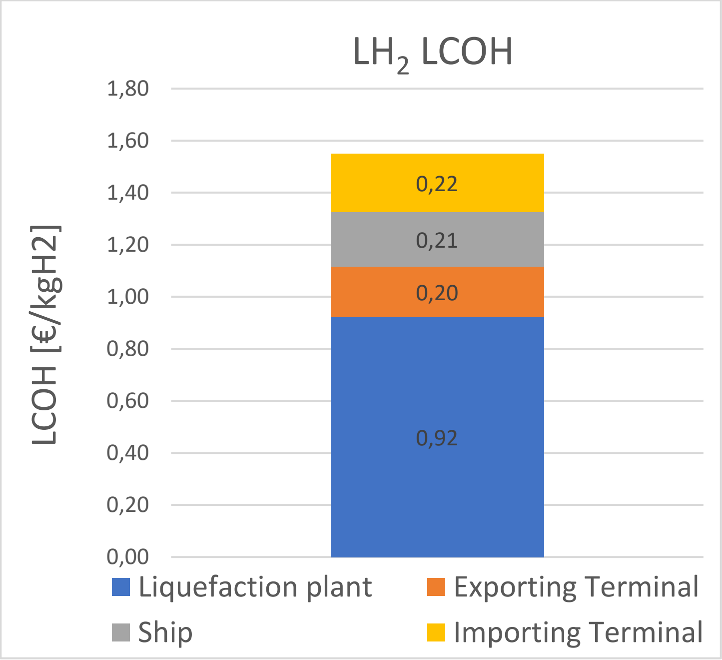


Figure 67 - LH2 LCOH break-through

5.2 Ammonia

Regarding the transport of hydrogen by the conversion in ammonia, the same reasonings made for the Liquid Hydrogen can be made since the daily hydrogen production and the distance of transport are fixed, the only parameter that can affect the LCOH is the number of ships. Therefore, once again, the LCOH as a function of the fleet is displayed in Figure 68:

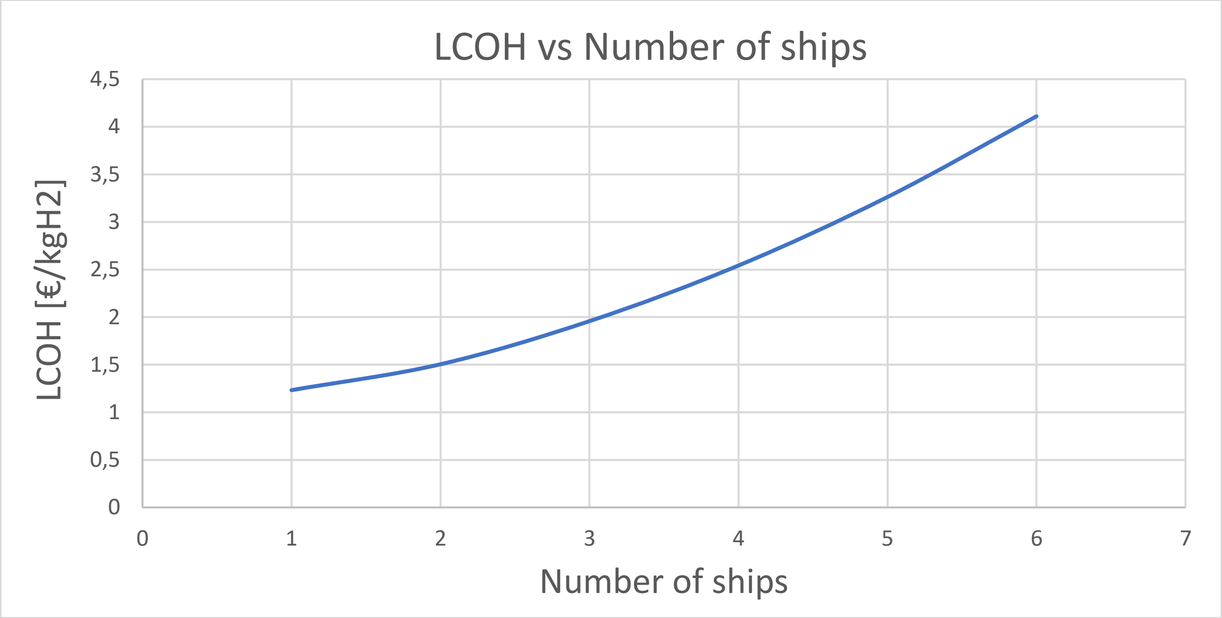


Figure 68 - NH3 LCOH depending on the number of ships

According to Figure 67, as it happened for the LH_2 case, the best choice to reach the lowest LCOH possible would be considering only one ship for the fleet. This configuration of the system leads to a final LCOH of 1,175 €/kgH_2 (Figure 69) lower than the LCOH of LH_2 (1,52 €/kgH_2).

The most influential parameters for the levelized cost of hydrogen, for the ammonia case, are the synthesis and cracking plants of ammonia. The synthesis accounts for 0,402 €/kgH_2, while the cracking facility for 0,379 €/kgH_2. This can be explained by both high capital and operative expenditures that such systems need.

Indeed, both the processes seen in section 3.4 and 3.8 (Ammonia synthesis and cracking) are highly energy demanding, even though not as much as the liquefaction plant needed by the LH_2 supply chain

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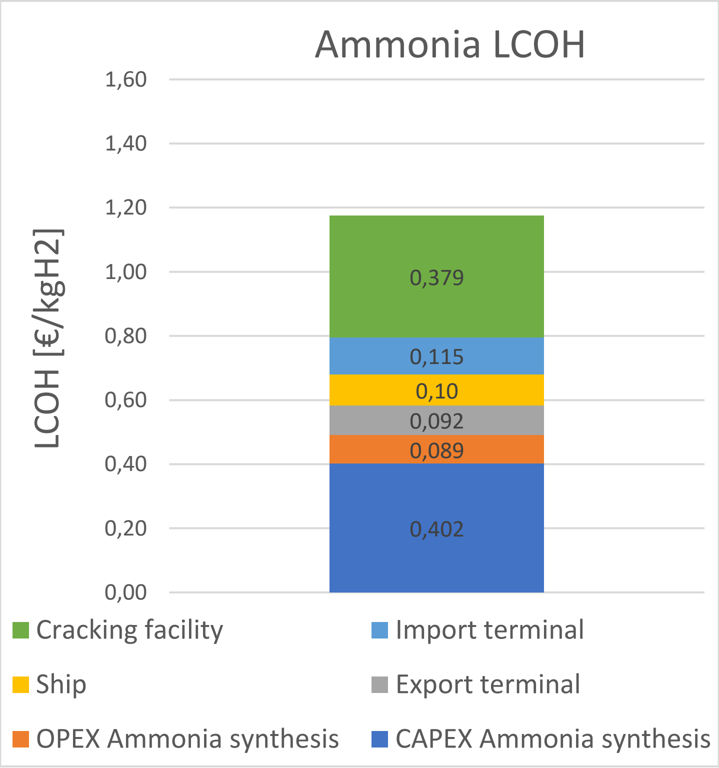

Figure 69 - NH3 LCOH break-through

5.3 Pipeline

In case of Hydrogen transport by pipelines, the distance of transport changes (from 1640km to 2000 km, of which 40km are underwater) because of natural boundaries that the shipping transport does not face. Unlike the case of LH_2 and NH_3, the parameter that has still influence on the LCOH, once the production rate and distance are fixed, is the distance among the compression stations. As already explained in section 4.6 with Figure 64, Figure 70 reports the best network design for the conditions previously declared.

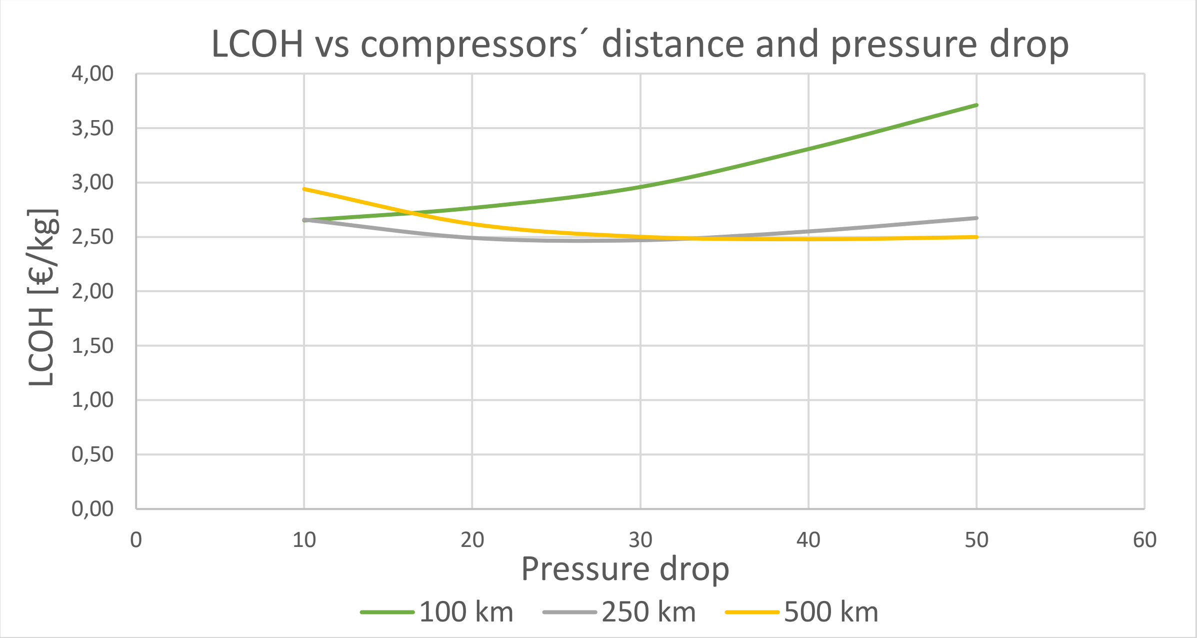

Figure 70 - Pipeline LCOH according to compressor distance and pressure drop

The difference in LCOH between a distance of 250 km and 500 km among the compression stations is negligible.

For the transport of hydrogen by pipelines, the optimal configuration shown in Figure 71 returns a LCOH equal to 2,47 €/kgH_2, way higher than the previous methods.

Indeed, only the Pipeline Capex alone (1,62 €/kgH_2) worth more than the others two ways of transport considered as a whole: 1,52 €/kgH_2 for LH_2 and 1,175 €/kgH_2 for NH_3.

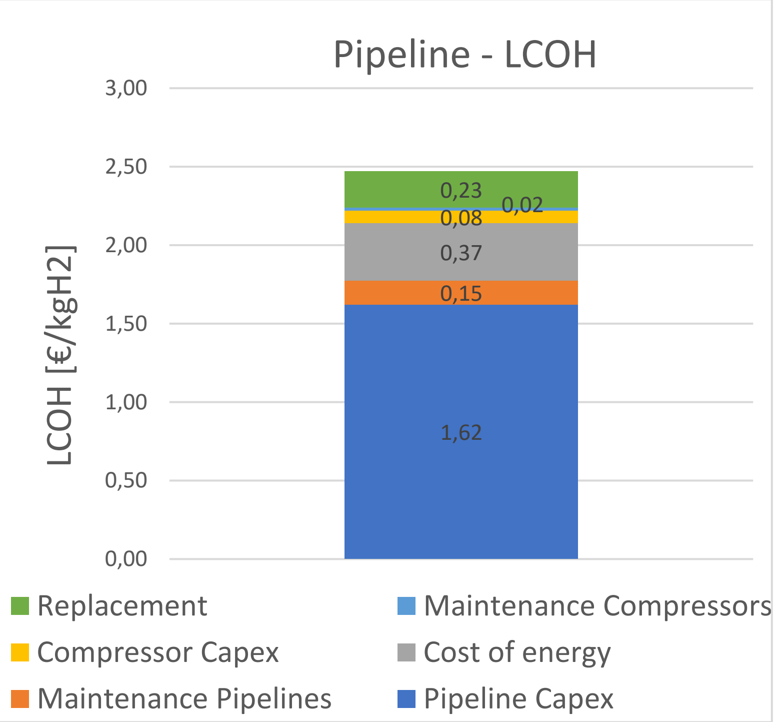

Figure 71 - Pipeline LCOH break-through

Results

In section 5.1, 5.2 and 5.3 the three ways of transport hydrogen (LH_2, NH_3 and Pipelines) have been analyzed, assuming a hypothetical case study where the distance of transport and rate of production were fixed according to the values reported in Table 15. The results of this analysis are summarized in Figure 72.

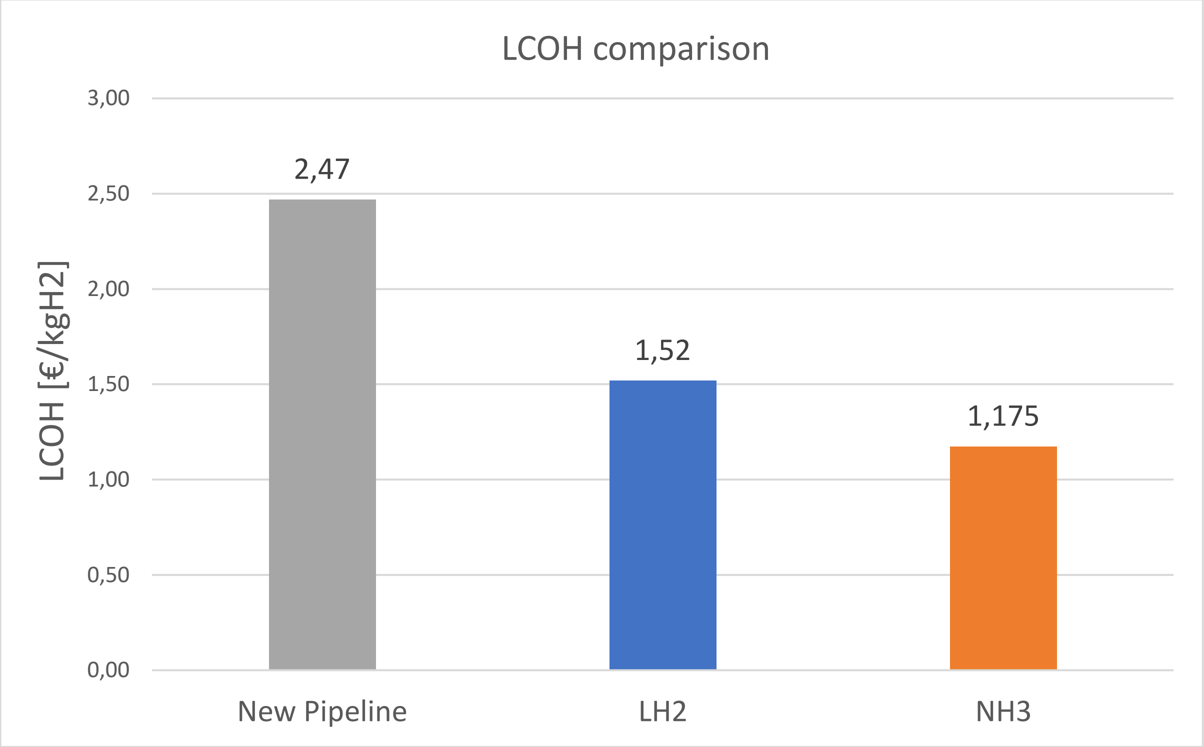

Figure 72 - LCOH comparison among the technologies

The best-case scenario indicates the transport of hydrogen by conversion in ammonia as best option available given such boundary conditions. The LCOH for a hypothetical transport of hydrogen converted into ammonia from Casablanca to Marseille, is equal to 1,175 €/kgH_2. This value is lower by 0,3 €/kgH_2 respect to the LH_2 transport, and less than half compared to the transport by pipelines. Such a large difference with the hydrogen pipelines can be explained by the daily quantity of hydrogen transported (500 tons/day). Indeed, pipelines are generally referred to as the best way of delivery energy for larger quantities of gas (up to 1500 or 2000 tons/day) and lower distance of transport (less than 1500 km).

Liquid Hydrogen and Ammonia are instead comparable. The most important parameter that can explain the difference in the LCOH is the large amount of energy required by the liquefaction of hydrogen; if the cost of energy in the future will decrease significantly, the LH_2 would take great benefit from it.

Nevertheless, all the results obtained for the 3 ways of transport are referring to the specific boundary conditions given in this case study, therefore would be a mistake assuming that ammonia is always the best way of transport hydrogen, regardless the specifications of the project. To further clarify this point, the following section will discuss the daily hydrogen production and transport distance ranges within each technology leads to the lowest cost.

Conclusions

Hydrogen can be a meaningful energy carrier in the energy transition according to private and government entities, helping decarbonize hard-to-abate sectors and representing a viable alternative to the electric vehicles for the sustainable mobility. Unfortunately, neither the production nor the transport supply chain is ready yet to face the expected rising demand by the market. New infrastructures and networks must be established among export and import countries. Indeed, the Levelized Cost Of Hydrogen will be sensibly influenced by the cost of electricity, needed by the electrolysis process and/or by any further conversion process the Hydrogen will undergo. The cost of electricity by renewable sources is then going to define new commercial routes or strengthening already existing commercial agreements.

Within this framework, this analysis sought to provide a benchmark on the estimation of hydrogen transport costs, spacing between different delivery scenarios and different transport alternatives (Figure 73).

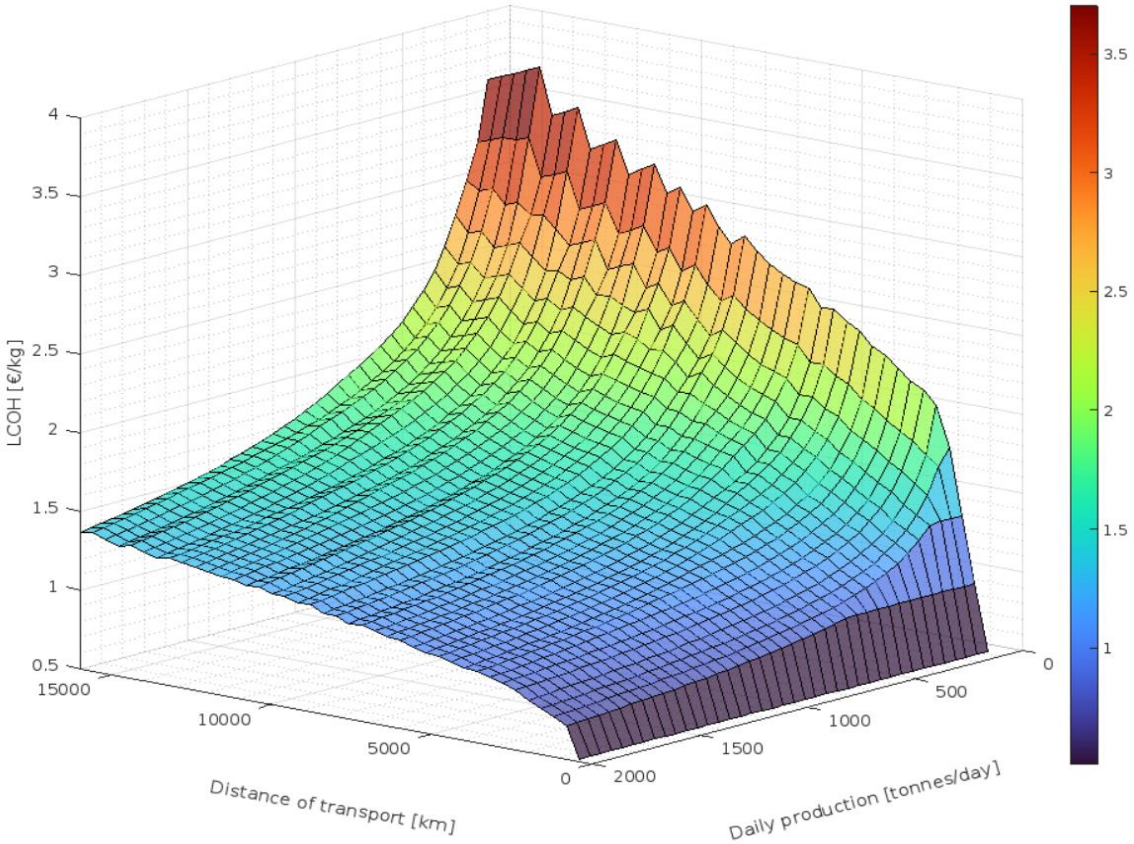

Figure 73 - Graphical representation of the minimum LCOH depending on production and distance of transport

The three technologies involved in the study (LH2, NH3 and hydrogen pipelines) have been analyzed regarding the economic and technical pillars of their systems, assigning cost functions to the most important cost-drivers, and estimating the energy consumption for the most energy-intensive processes. Doing so, a dynamic tool able to return the value of LCOH for each technology has been developed. The ultimate goal of this work was, given the boundary conditions of a project, to highlight not only the hypothetical cost in €/kgH_2 (Figure 73), but also the best technology according to the project specifications (Figure 74).

From Figure 73, we can assert that when the Hydrogen production increases, the whole system benefits from it, lowering the specific cost of Hydrogen (in €/kgH_2). On the other hand, increasing the distance of transport will bring an inevitable jump in the specific cost of delivery. Higher specific costs of transportation into the extreme values of the graph (3,62 €/kgH_2), when the production is equal to the minimum value, 100 tons/day and the distance of transport is set to the maximum value, 16000 km.

Nevertheless, the information missing in Figure 73 is the best technology to be adopted to guarantee the lowest LCOH possible shown. This information can be graphically seen in Figure 74, where a color has been assigned to each technology: Blue for Liquid Ammonia, Yellow for Pipelines and Green for Liquid Hydrogen.

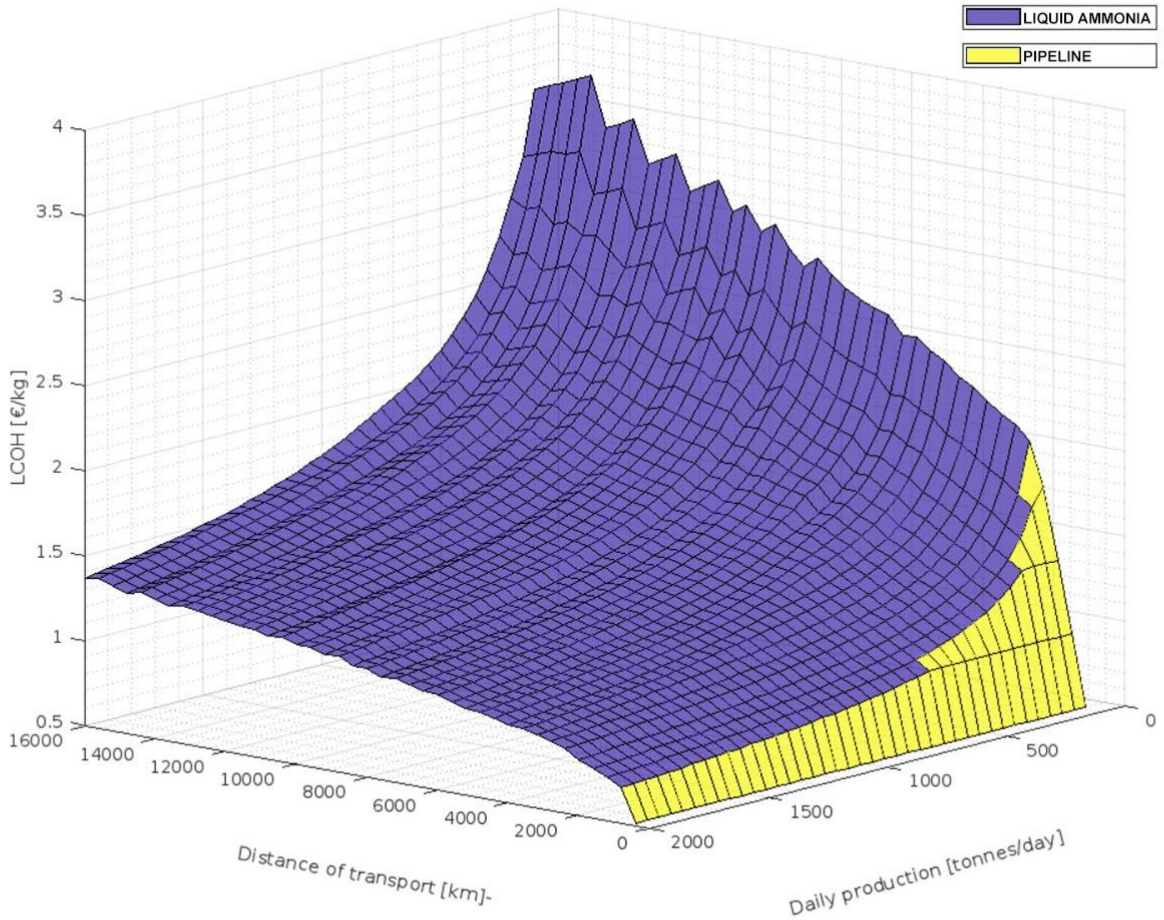

Figure 74 - Graphical representation of the minimum LCOH according to the best technology

The green color that stands for Liquid Hydrogen, does not appear in any area of the graph, because is always hidden by the other two ways of transport. Yet even though Liquid Hydrogen is not shown, for medium distances (between 500 and 5000 km) and high amount of hydrogen delivered (between 500 and 2000), is very close to compete with ammonia. Nevertheless, these last results must be carefully checked in the next years; there are still many uncertainties about the real costs of the ammonia cracking facilities and depending on the chosen scenario ammonia could be even more expensive than LH2. On this regard an example of how Figure 74 would change in case of a pessimistic scenario for the Ammonia cracking costs, is shown at the end of Appendix III.

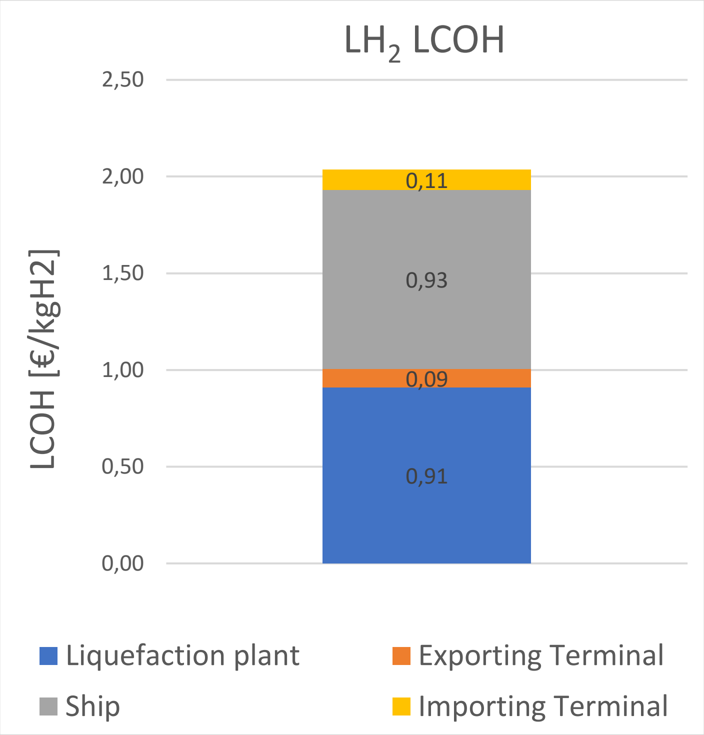

Figure 75 - LH2 LCOH break-through

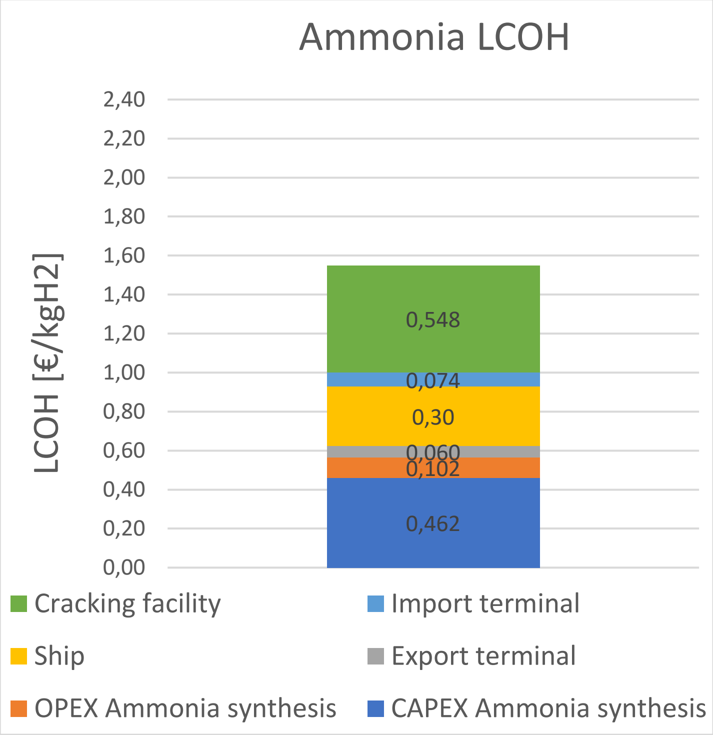

Figure 76 - NH3 LCOH break-through

For higher amounts and shorter distances, Hydrogen pipelines are by far the cheapest solution, while for longer distance, the BOG losses limit the efficiency of the LH_2 supply chain.

However, the main reason behind the LH_2 transport lack of competitiveness relies on the OPEX and CAPEX of the Liquefaction plant, which is the main cost driver of such technology.

As already discussed previously, the Liquefaction plant alone is worth more (0,92 €/kgH_2) than the Ammonia synthesis and cracking plants investment combined (0,402€/kgH_2 + 0,089 €/kgH_2 + 0,379 €/kgH_2).

The two LCOH (Figure 75, Figure 76$) for Ammonia and Liquid Hydrogen found for the case study are reported as follow for sake of easier graphical visualization.

Finally, the transport of Hydrogen cannot be dominated by a single specific technology, but rather must be open to different design choice that would guarantee the optimal LCOH.

Higher amounts of Hydrogen transported for short distances are unambiguously cheaper when delivered by pipelines, while ammonia has a larger range of operation.

Liquid Hydrogen could compete with the previous two way of transport if the Capex of the liquefaction plant will be more affordable and if the forecasts for the energy consumption will be respected.

Appendix I – Liquid Hydrogen

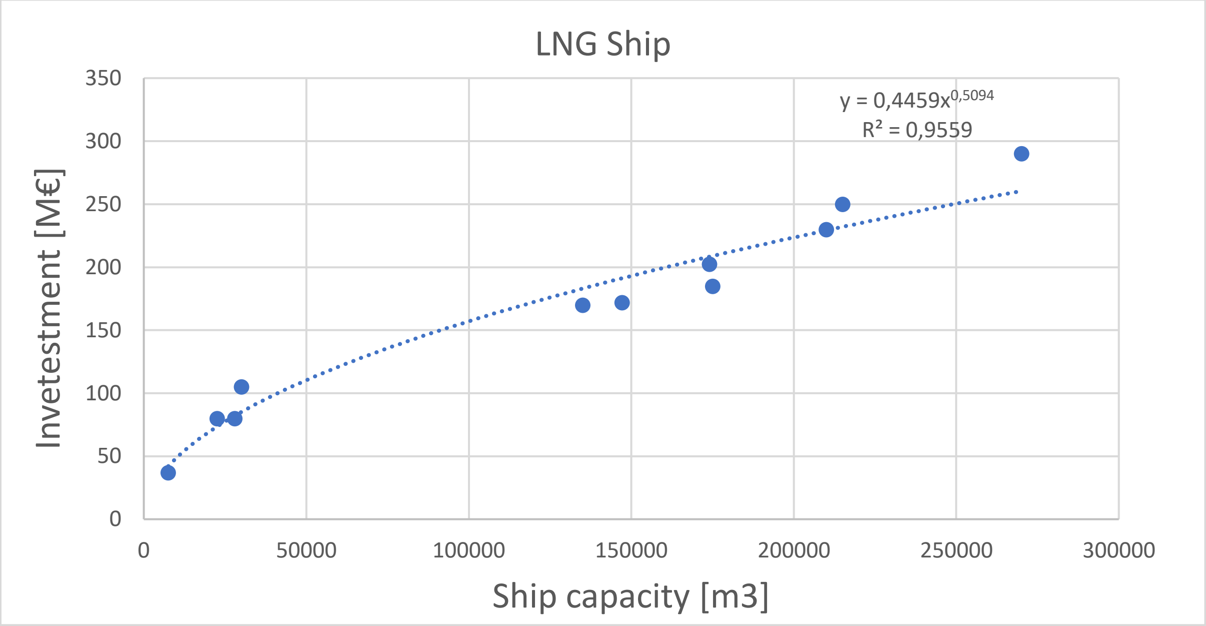

Figure 77- LNG ship investment cost

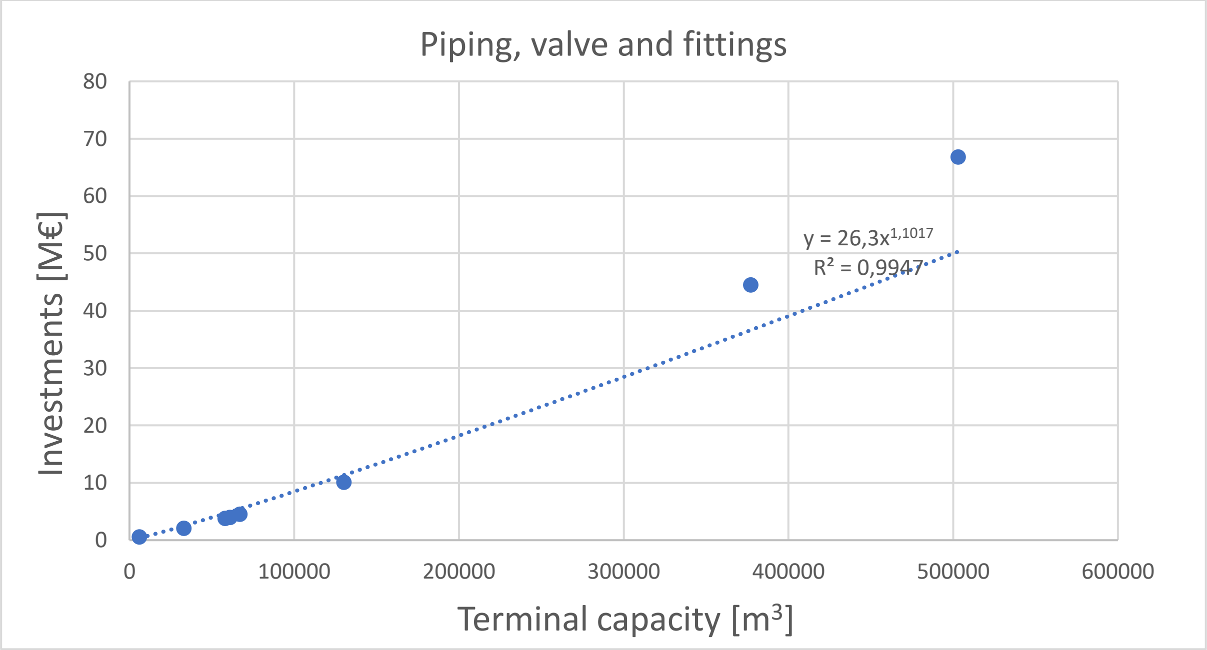

Figure 78 - - Piping, valve and fitting investment according to the terminal capacity

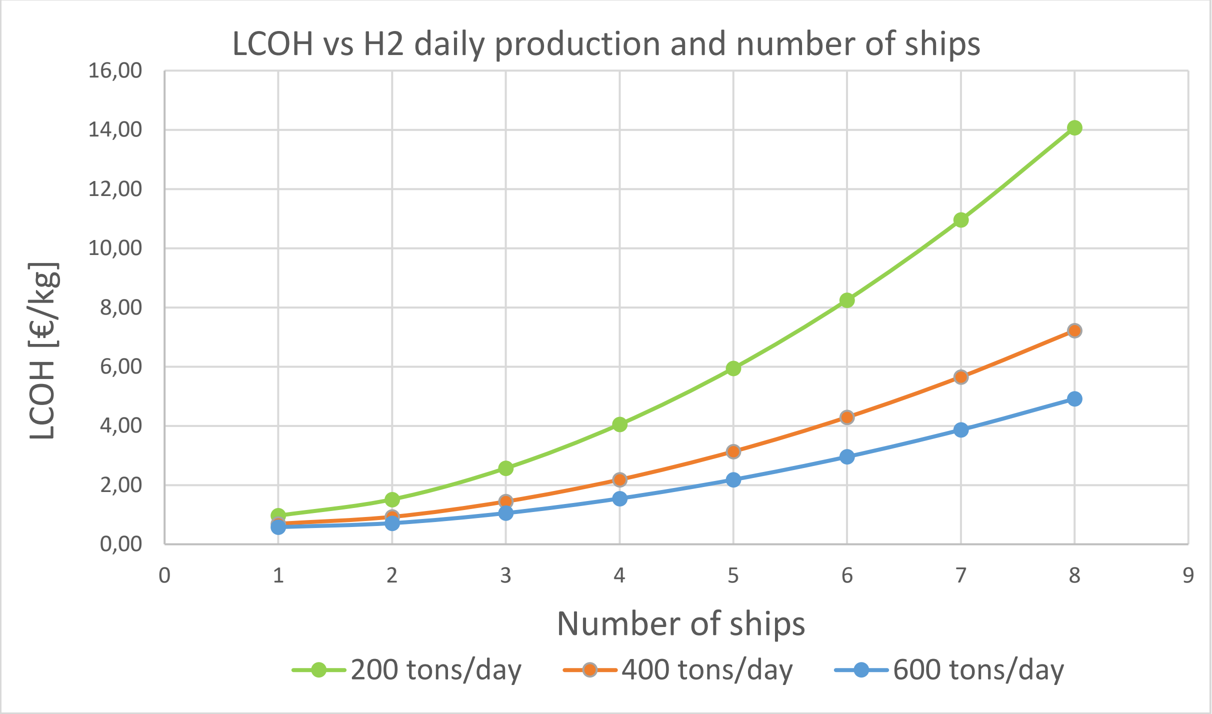

Figure 79 - LCOH of LH2 vs daily production and number of ships

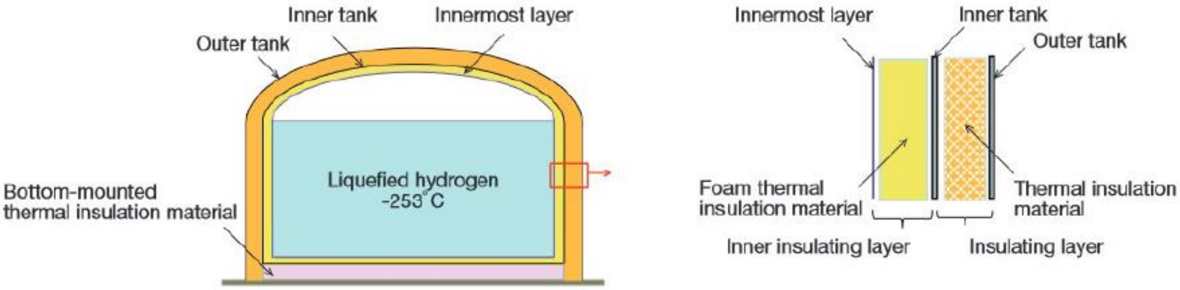

Figure 80 - Future large-scale liquid hydrogen storage - image from [17]

Appendix II – Ammonia

On the following Figure 81 is possible to see how the fleet investment changes with regard to ships´ number. Fixed the annual delivery of ammonia, for the highest number of ships take in consideration in the analysis the overall investment is equal to over 330 millions of euro, while for the one ship case the investment is slightly more than 130 millions of euro.

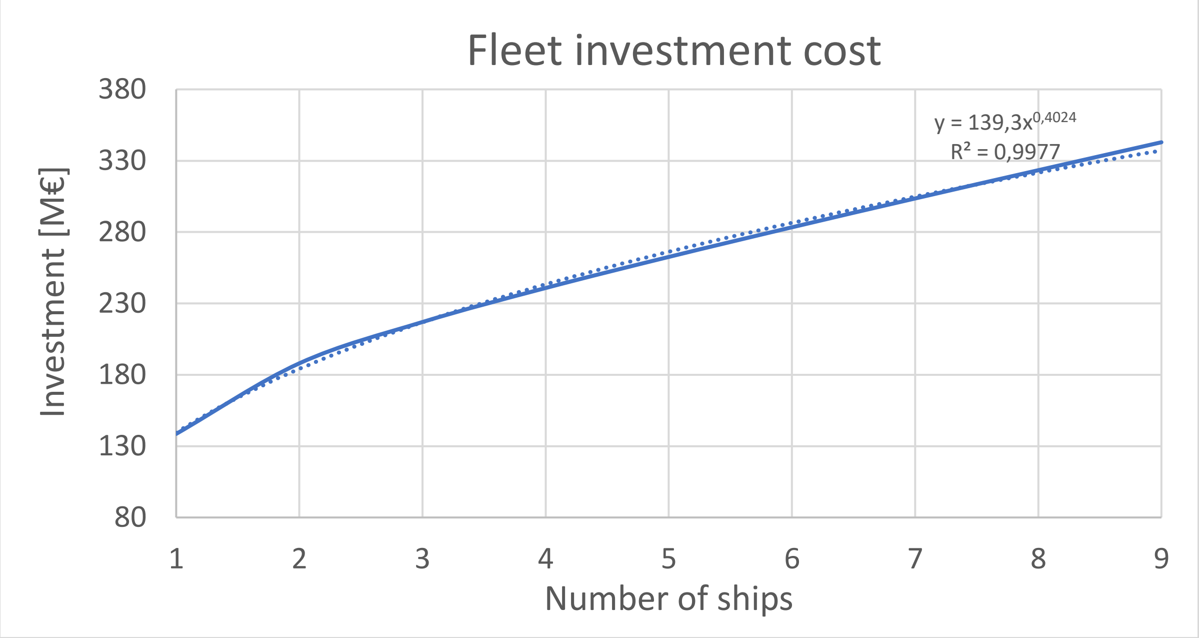

Figure 81 - Fleet investment cost with respect to the number of ships

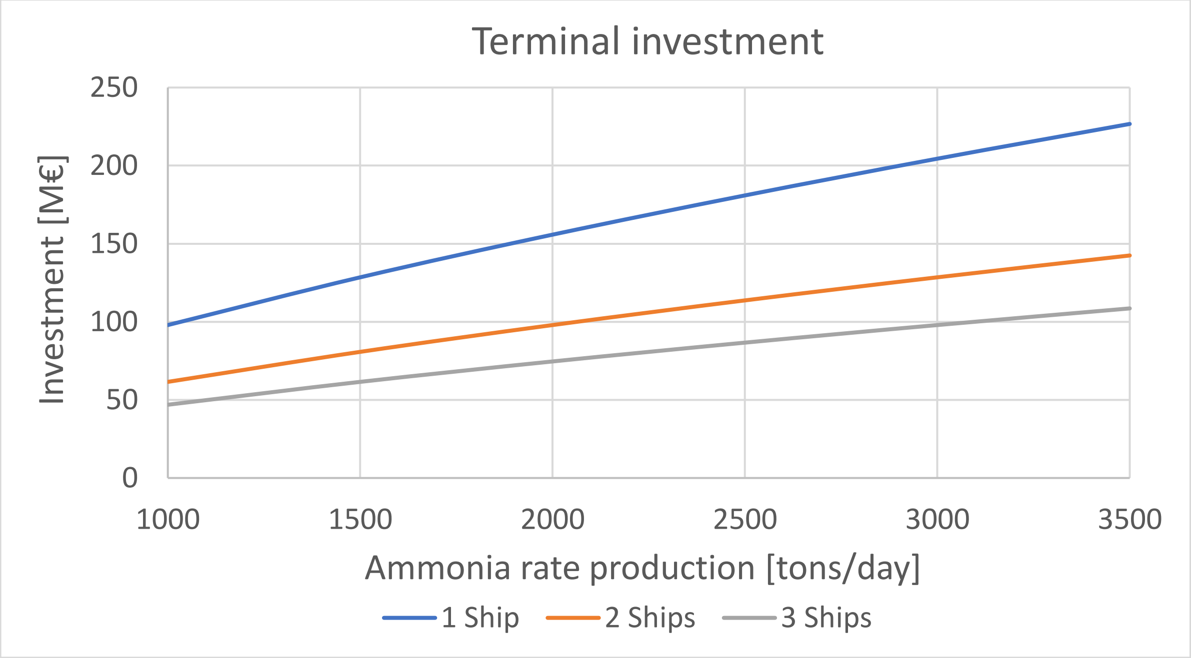

Figure 82- Terminal investment vs different number of the ships assuming different ammonia daily production

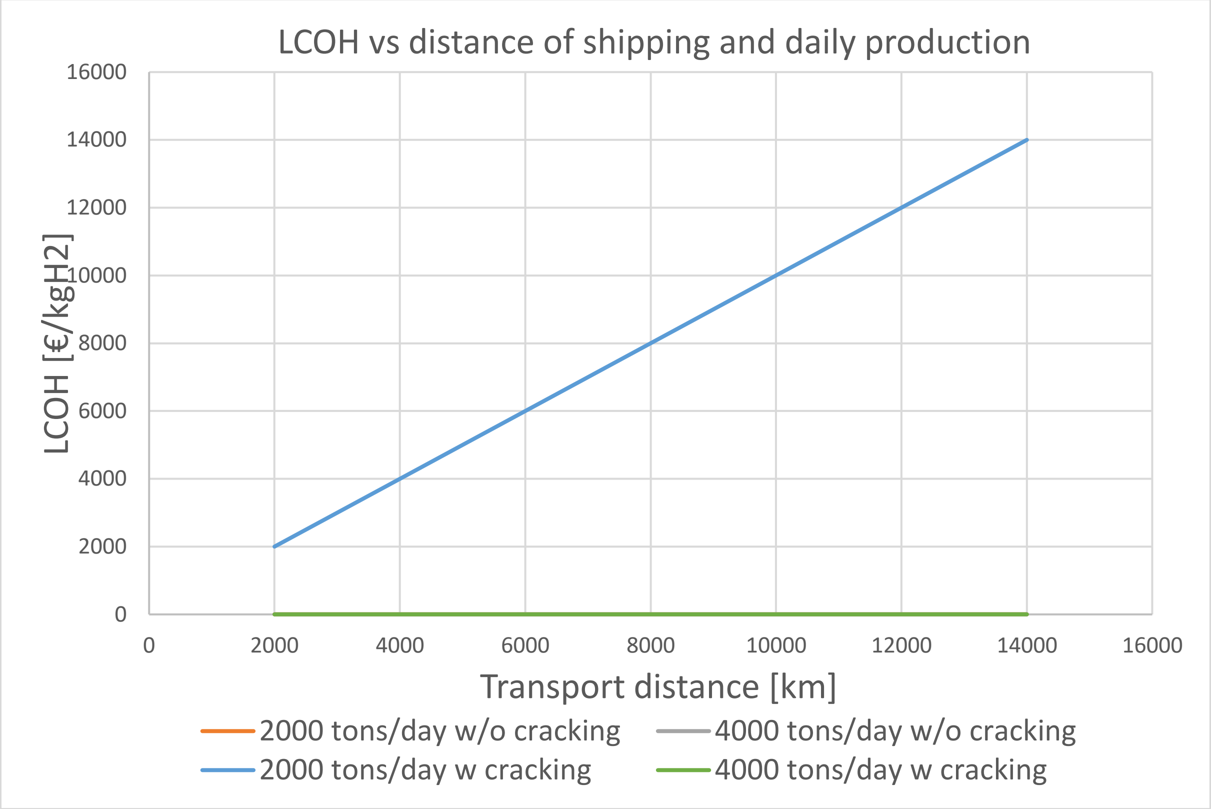

Figure 83- LCOH vs transport distance and daily production

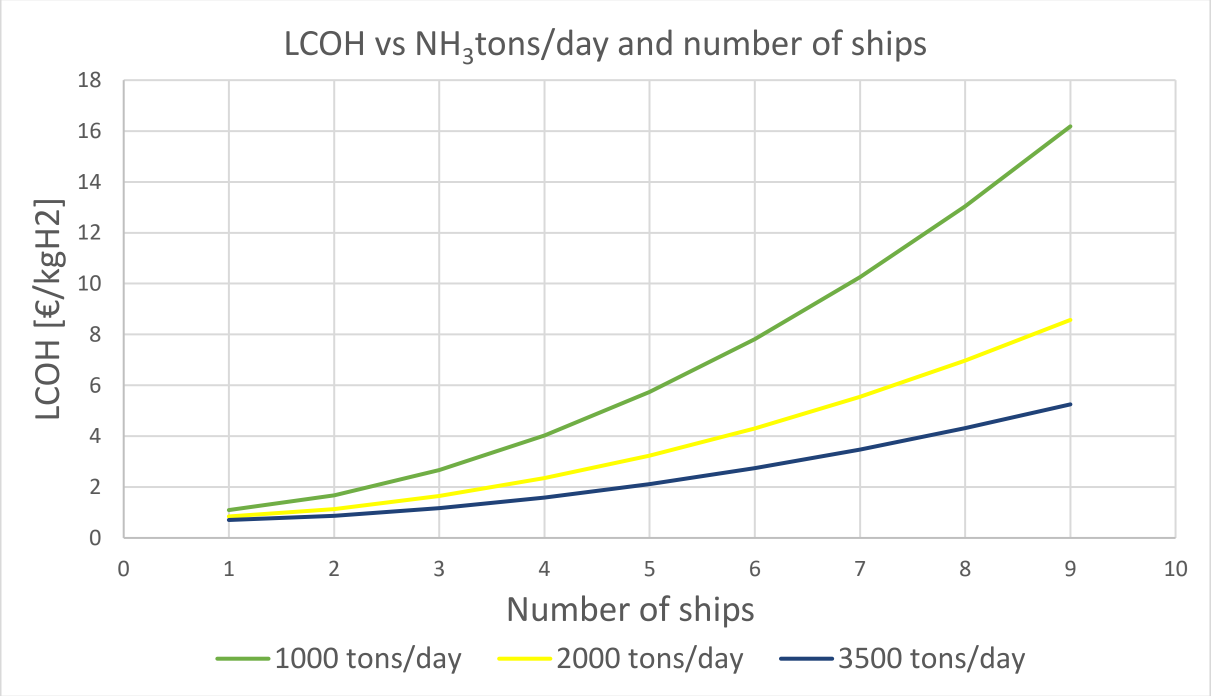

Figure 84- LCOH vs NH3 tons/day and daily production

Appendix III – Pipelines

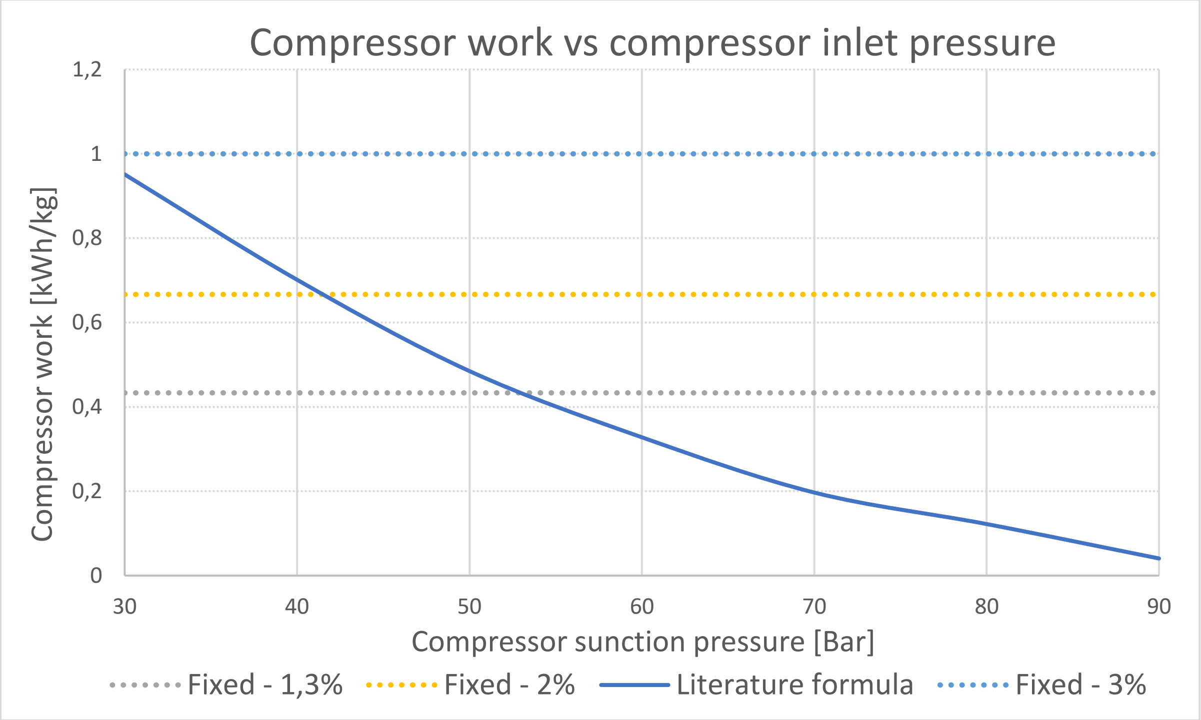

Figure 85 - Literature equation for compression work

Black line [53] :

$$P_{multi\ stage} = N\left(\frac{k}{k-1}\right)\left(\frac{Z}{\eta_{isen}}\right)T_{suc}(q_M)R\left[\left(\frac{P_{disc}}{P_{suc}}\right)^{\left(\frac{k-1}{Nk}\right)}-1\right]$$

(3.2) - Compression

Blue line [62] :

$$BHP = 0,653 \cdot Z_{ave}\left[\frac{(Q_n)(T_1)}{E\cdot\eta}\right]\left[\frac{\kappa}{\kappa-1}\right]\left[\left(\frac{p_2}{p_1}\right)^{\frac{\kappa-1}{\kappa}}-1\right]$$

(3.3) - Compression

Greenline: [63]

$$P = \frac{Q}{3600*24*33.33}*\frac{ZTR}{M_{H_2}\,\eta_{comp}}*\frac{N\gamma}{\gamma-1}*\left(\left(\frac{P_{out}}{P_{in}}\right)^{\frac{\gamma-1}{N\gamma}}-1\right)$$

(3.4) - Compression

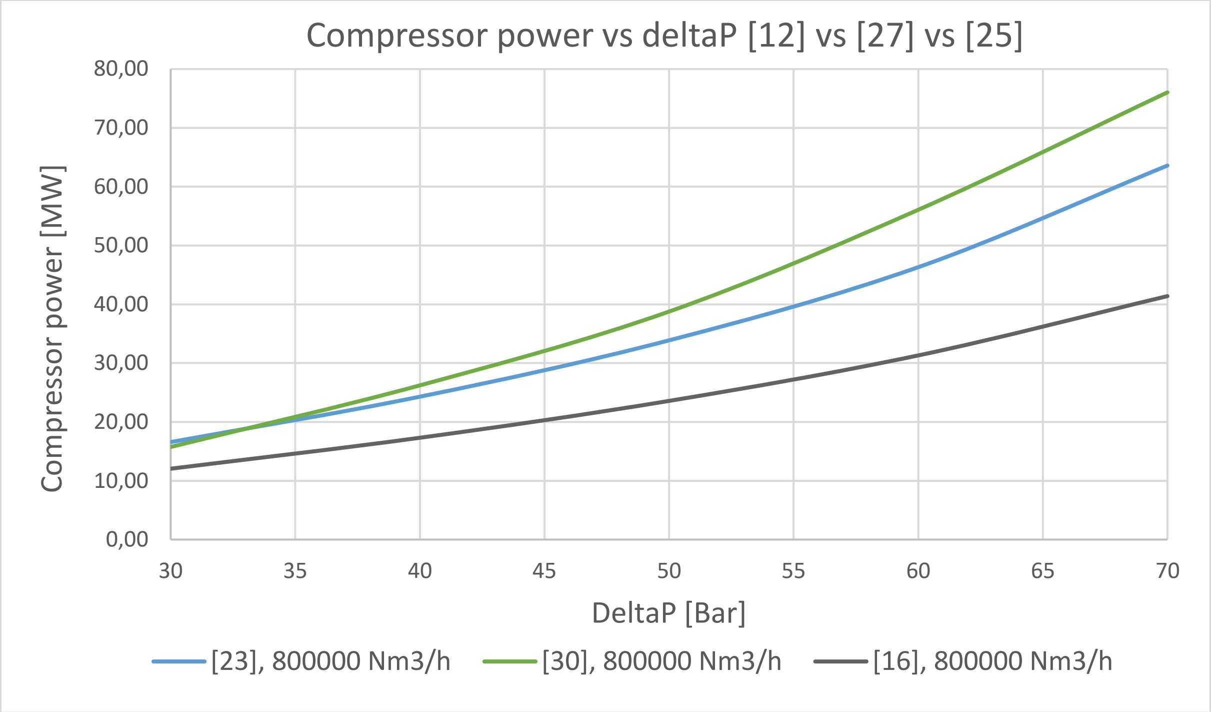

Figure 86 - Compression power for different equation available on the literature

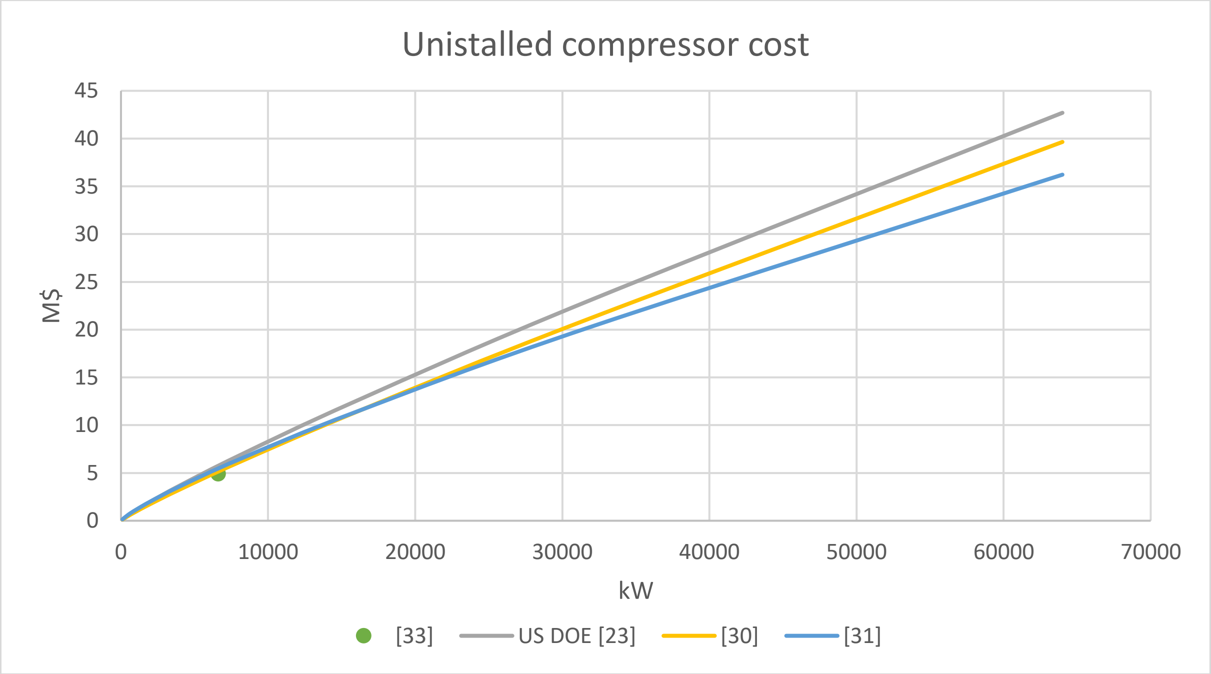

Figure 87 - Uninstalled compressor cost; different literature approaches

Yellow line: [64] Grey line: [26] Blue line: [65] Green dot: [66]

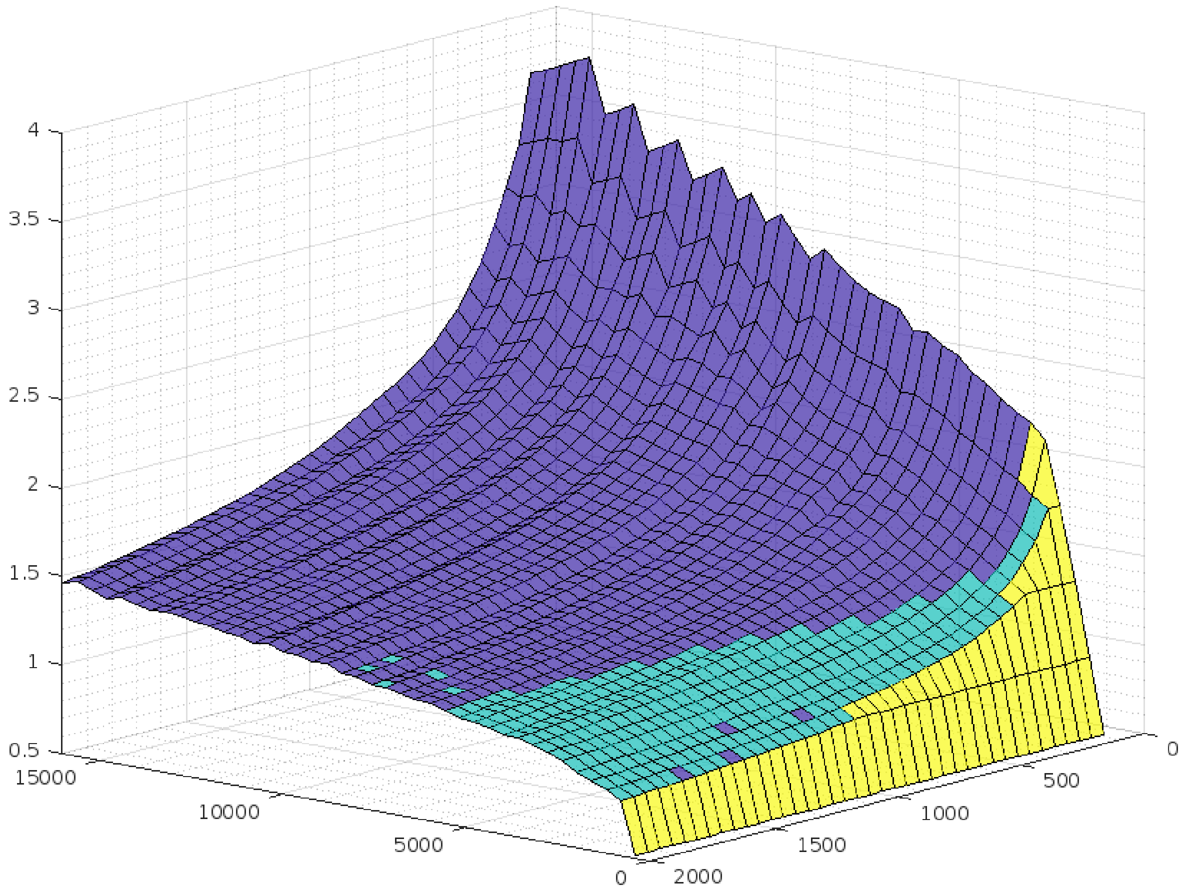

Figure 88 - Graphical representation of the minimum LCOH according to the best technology - Pessimistic Ammonia scenario

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