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Techno-economic Analysis of Cryogenic Deblending Installations for Gas Quality Management in Hydrogen-blended Gas Networks

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Abstract

This report attempts to provide useful information in the context of hydrogen blended with methane distribution using the National Gas Distribution currently transmitting and distributing Methane.

Since the need to deblend back to Hydrogen or Methane to meet customers' needs is inherently coupled with this blending distribution scenario, this report provides technoeconomic figures and characteristics of this deblending facility. It individualizes the characteristics of the mentioned equipment at any given point on the grid where the facility is required.

The main input parameters are the mass flow of the blend and its inlet pressure and delivering pressure (required) mainly at points where PRI (Pressure Reduction Installation) are present.

A model has been carried out to simulate a real cryogenic gas separation process. Subsequently, the components and internal variables have been optimized and this has shown that the configuration is able to achieve cryogenic temperature leveraging from the pressure gradient involved in the PRI enabling the hydrogen to be recovered from the blend at high purity >98%.

Additionally, the technoeconomic analysis has reinforced the feasibility of the deblending facility with LCOH's (levelized costs of kilogram of hydrogen deblended) fluctuating within the range of 0.158 to 0.65 €/kg.

A sensibility analysis has been run on to the main variables that define an optimum solution at a given PRI showing the existence of a tradeoff between the HXs area and the recompression power required.

Likewise, another sensibility analysis has been run over the LCOH by evaluating different PRI pressure drops showing the LCOH to be very sensitive to higher pressure delivery targets. This is because the LCOH increases sharply due to higher compression power required to meet the high delivery pressures. Finally, further LCOH sensitivity analyses were trailed showing the LCOH to be insensitive to both the hours of operation of the deblending machine (yearly) and to the price of electricity.

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E. List of abbreviations

Mt	Millions of tones
BCM	Billion Cubic Meters
IEA	International Energy Agency
PRI	Pression Reduction Installation
NTG	National Transmission Gas
NG	Natural Gas
LNG	Liquified Natural Gas
J- T	Joule Thomson (expansion)
WGS	Water gas shift (reaction)
ORC	Organic Rankine cycle
GB	Great Britain
HX's	Heat Exchangers
SV	Separator Vessel
MCMD	Million cubic meter day
H2	Hydrogen
LP	Low Pressure
MP	Medium Pressure
HP	High Pressure
CH4	Methane
LCOH	Levelized Cost of Hydrogen mass (€/kg)
LCOE	Levelized Cost of Energy (€/MWh)
HPRV	High Pressure Reduction Valve
MPRV	Medium Pressure Reduction Valve
LPRV	Low Pressure Reduction Valve
HPRV	Hydrogen Pressure Reduction Valve

Chapter 1 Introduction and Objectives

1.1 Introduction

In 2021 European hydrogen producers generated 75 MtH₂ as pure hydrogen and an additional 45 MtH₂ as part of a mix of gases of hydrogen (IRENA), which covered a total demand that reached 94 Mt that year (IEA). Under all projected scenarios towards meeting decarbonization targets ranging from the boldest to the most conservative considerable increases in hydrogen use is found which is destined to increase to reach around 500 Mt by 2050. This is expected to take place in the following years.

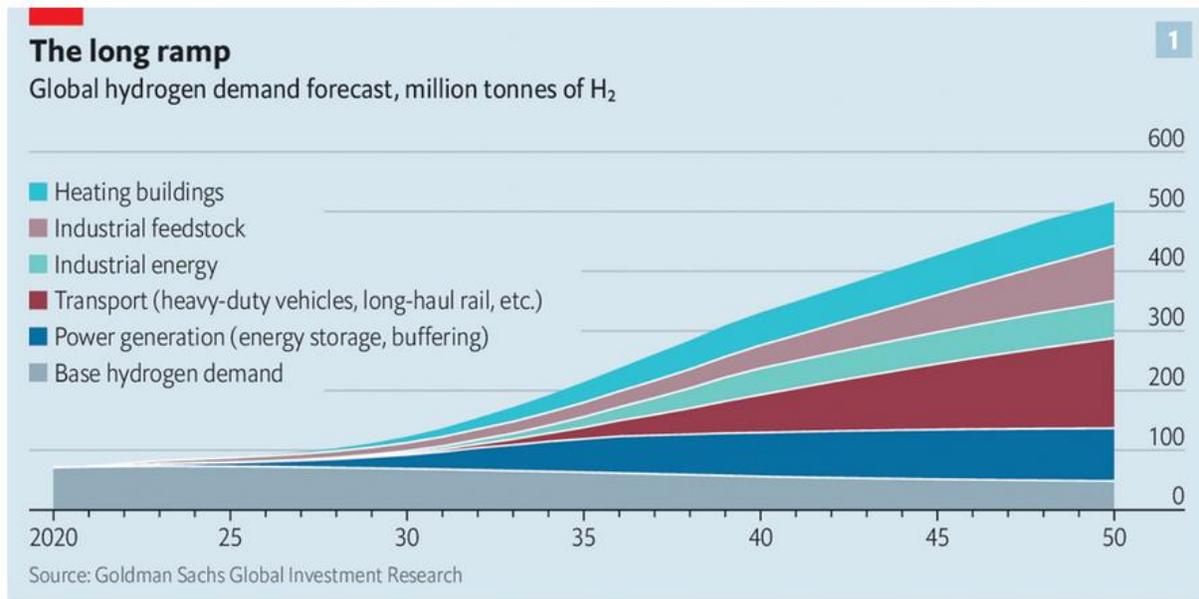


Figure 1 Global hydrogen demand forecast (Economist, 2021)

This trend is also stated to take place by the IEA World Energy Outlook 2022:

“Concerns about fuel prices, energy security and emissions – bolstered by stronger policy support – are brightening the prospects for many low-emissions fuels. Investment in low-emissions gases is set to rise sharply in the coming years. In the APS, global low-emissions hydrogen production rises from very low levels today to reach over 30 million tonnes (Mt) per year in 2030, equivalent to over 100 BCM of natural gas (although not all low-emissions hydrogen would replace natural gas). Therefore, methane is an actor that continues to be in the energy matrix composition along with the incorporation of alternative fuels. Renewable and low-carbon gases are indispensable to fully decarbonize the European Union (EU) energy system (Carmen Wouters, 2020).

Hydrogen is one of those renewable gases with much of it being produced close to the point of use, but there is growing momentum behind international trade in hydrogen and hydrogen-based fuels. Projects representing a potential 12 Mt of export capacity are in various stages of planning, although these are more numerous and more advanced than corresponding projects to underpin import infrastructure and demand.” (IEA, World Energy Outlook 2022 - Executive summary, 2022).

Therefore, the import infrastructure to distribute domestically that traded hydrogen and the demand for it is lagging behind.

1.1.1 Blending hydrogen with methane utilizing existing gas distribution infrastructure and operation.

Under these circumstances endeavors that enable the growing in the import infrastructure are needed. One of these endeavors that have gained interest in this line is the integration of new energy vectors (like hydrogen) into the already existing gas infrastructure by blending it with Natural Gas. This idea makes sense when acknowledging that transmission lines able to distribute pure hydrogen need to be heavily retrofitted or made brand new to handle 100% hydrogen gas so that cracking, leaking and operative problems such as pressure reduction are avoided.

Injection of hydrogen in the natural gas transmission system could be a smart way to utilize the grid to transmit and distribute a mixture of blended hydrogen and natural gas in an agreed mole fraction ration between the two and is a use case that is currently being considered in some countries as it entails a viable and gradual transition towards 100% hydrogen transmission and distribution instead of something that happens from one day to the next.

This brings economic benefits due to the already acquired know-how knowledge of natural gas distribution and operation and the avoidance of having to construct a new infrastructure of a distribution system from scratch.

Basically, the concept of a central operator is kept as in the current NTG system but it is added the need of 1) managing the grid levels of the blend assuring that the molar fraction of the two gases are kept to previously agreed values; and 2) modify the lines and equipment so they can endure a different gas composition; this is especially relevant when hydrogen starts to become prominent in the blend as it is observed that as the hydrogen molar fraction in relation to NG begin to increase the more robust the pipelines as well as the operative equipment (presence of pressures stations) need to be.

The technicalities and the complexity change (the mentioned elements are not minor) but the deployment and operation of the gas distribution system managed in a central way is kept.

Provided strong policy is backing this up, this system represents a clear and cost-effective opportunity to cut emission concordantly with climate change goals. It pushes forward the energy transition since once in place encourages users to acquire appliances that run on hydrogen such as fuel cells or machinery (turbines) adapted to combust hydrogen (given that there is availability of hydrogen). At the beginning there might be also demand for pure methane which will have to be met too (for hydrogen sensitive users), but a trend that advances on higher hydrogen adaptation is expected to come with the decision.

In this context, the need for additional deployment of deblending equipment installed downstream the grid distribution (PRI and offtakes) able to separate the hydrogen back from the blend is needed (also to separate the methane from the blend for hydrogen sensitive users). The pressure gradient existent throughout the grid at different operating points can be leveraged as an energy driver to run the separation process.

1.1.2 State of the art of Gas Separation equipment related to hydrogen recovery.

The knowledge on gas separation (deblending) has been developed and refined considerably and the technology associated also is already mature. Particularly, the hydrogen industry, without going any further, uses gas separation mechanisms to produce hydrogen from Natural gas, an industry that traces back up to the early 70's. The well-known standard method is called Water Shift Reaction (WSR) which apart

from methane uses water steam. Further down the process stages the gas results is Syngas which then needs to undergo a gas separation process to recover the hydrogen.

In general, for any application diverse gas separation methods can be used with the main being cryogenic separation, membrane separation, pressure swing absorption or a combination of these.

Among these we are supposedly interest in analyzing the cryogenic separation method that have already been proven valid. In particular models for treatment and hydrogen purification through cryogenic methods using different software's and later implementation have been carried out.

1.1.3 Existing Piping use feasibility

In subsequent stages all the points subjected to be chosen to accommodate deblending capacity need to be checked further based on whether it is feasible to retrofit that NTG infrastructure to endure the specific gas composition and mass flow and then in the case that the retrofitting is feasible determine the extent of the costs of the initial work implicated to incorporate them as a basis to develop more detailed evaluations in consideration of site-specific conditions and scenario for hydrogen concentration with increasing hydrogen production and blending (National, 2021).

1.2 Goals and Objectives

1.2.1 General objectives

Provide a thorough overview of the project of using the NG grid for blending hydrogen and methane. Describe the integration of the systems that need to be in place for this option to be real, so it contextualizes its feasibility.

1.2.2 Specific objectives

The specific goal of this report is to model of a cryogenic gas separator for deploying in the context of a blending hydrogen with methane into NGT system scenario in which the deblender facility is place at the PRI (Pressure reduction installation) that are in between the Transmission and Distribution Grids. The model therefore simulates a deblender that will produce hydrogen at this point and so it will deliver the equipment's characteristics and performance of it. The Italian case will be used as it is of special interest. For which a dataset of operational conditions such as the mass flow, inlet and outlet pressure provided from the Italian grid operator are used to evaluate real relevant cases for Italian grid.

Chapter 2 Literature Review

2.1 Technoeconomic evaluation feasibility of deblending facility deployment in GB's NTG.

Deblending (i.e. separation of the blended gas stream) is a potential solution to allow the existing gas transmission and distribution network infrastructure to transport energy as a blended gas stream (National, 2021)¹.

The technologies noted above require an energy input to drive the separation of gas components in the form of compression to provide a pressure differential. The configuration of the GB gas transmission and distribution networks provides a possible source of available energy through the pressure let-down in the network pressure tiers, which could be used to drive the gas component separation processes (National, 2021).

An assessment has been carried out in the GB Gas Grid network operated by Cadent and Northern Gas Network (NGN) considering as potential points for deblending.

NTS offtakes, where gas is feeding to Gas distribution networks (GDN) or industrial/power plants and PRIs, where the gas pressure is diminished by local transmission systems for delivery. Through scattering realistic gas network operating conditions at these points, representative operating pressures are used as representative basis in order to carry out the technoeconomic analysis.

The representative base withdrawn from the scattered measures of the operating pressures conducted for both offtakes (Local transmission systems and Industrial users) and PRI is the following:

Table 1 Basis for Blended Gas feed (National, 2021) ²

Parameter	1A	1B	2A	2B
Network Location	LTS PRI	LTS PRI	NTS offtake	NTS offtake
Flow (million Sm ³ /day, mcmd)	1	1	3	3
Feed Gas Inlet Pressure (barg)	30	30	60	60
Residue Gas Outlet Pressure (barg)	2	20	7	30
Hydrogen Product Pressure (barg)	20	20	20	20

- In this work they come up with the separator process performance linked to each one of these base use cases.
- This is carried out for two scenarios namely the one using Cryogenic Separation technology and the one using Separation Membrane.
- The base case considers production of combustible hydrogen, with a purity of >98 mol% hydrogen and <1 mol% hydrocarbons and CO₂ (National, 2021).

¹ Note that it might play a role not only in providing pure hydrogen or pure methane to sensible customers but also in balancing (maintaining) the grid blended mole fraction.

² For the hydrogen product, the pressure is 20 barg nominal for all cases, chosen to allow a consistent basis for comparison between the technologies evaluated. 20 barg is an indicative operating pressure.

- In both cases analysis it is assumed that hydrogen product demand matches the hydrogen production of the debinding facility. In other words, there are no constraints in terms of minimum or maximum demand of hydrogen product which would impact the scope of facilities and costs.
- The reference hydrogen content in gas blend to be considered as the base case is 20 mol% hydrogen.
- Additionally, 5%, 10% and 40% have been considered to observe the implications for hydrogen debinding for a range of likely hydrogen content in feed gas.

The process is generally carried out in a multistage arrangement, each stage operating at a reduced temperature as shown in Figure 2.

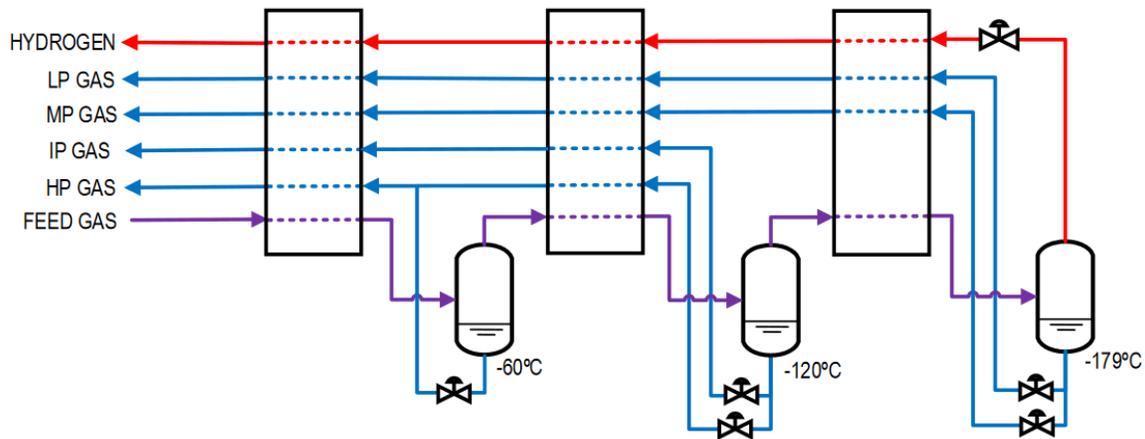


Figure 2 Multi stage cryogenic (Low temperature) Separation process. (National, 2021)

2.2 Cryogenic Cycles

The term cryogenics is used to describe methods of refrigeration at very low temperatures (typically below $-148\text{ }^{\circ}\text{C}$) and distinguish them from ordinary refrigeration cycles. Many of these methods relate to the liquefaction of gases known as permanent, like air, natural gas, hydrogen or helium.

Cryogenics is a field in engineering that focuses on systems operating at very low temperatures and which poses special problems, particularly in terms of fluids and materials.

Refrigeration and cryogenic liquefaction cycles involve combinations of isothermal compressions, cooling, thermal regeneration, and isenthalpic or adiabatic expansion of fluids.

There are three major families of cryogenic thermodynamic processes:

- a) isenthalpic expansion Joule-Thomson processes.
- b) isentropic expansion reverse Brayton cycles.
- c) mixed processes involving isenthalpic and isentropic expansion (Claude cycle).

Following we illustrate and explain these 3 cycles applied to liquefaction cases to better understand and relate the process aimed at this work of recuperation of hydrogen from a cooled mixture of H_2 in gas phase and methane in liquid phase.

2.3 Joule-Thomson Isenthalpic Expansion Process

These cycles are composed of valves (or a turbine), compressors and external refrigeration units and use isenthalpic expansion that have two drawbacks: firstly, the expansion work is lost, and secondly cooling

cannot be achieved if the fluid thermodynamic state is such that the Joule Thomson expansion leads to a temperature increasing.

2.3.1 Methane liquification basic cycle

An example of the J-T Isenthalpic expansion Process used to produce Liquefied Natural Gas (LNG) is considered below.

To liquify natural gas, methane is taken at 1 bar and 7 °C and is compressed to 100 bar to then be cooled to -63 °C (it is assumed in this example that an external refrigeration cycle is available for that).

The very high compression ratio (1:100) requires the use of several compressors (3 in this example) even though compression is assumed to be isentropic. Between compressors intercooling is provided to maintain the 7 °C of the feed methane. Intermediate pressures are equal to 5 and 25 bar.

Then the gas cooled to -63 °C and it is isenthalpically (use of valve) expanded from 100 bar to 1 bar by which gas and liquid phases are separated (J-T cooling principle).

As shown below (Figure 3), the methane enters the system, and liquid methane (product) and gaseous fraction exit from the separator.

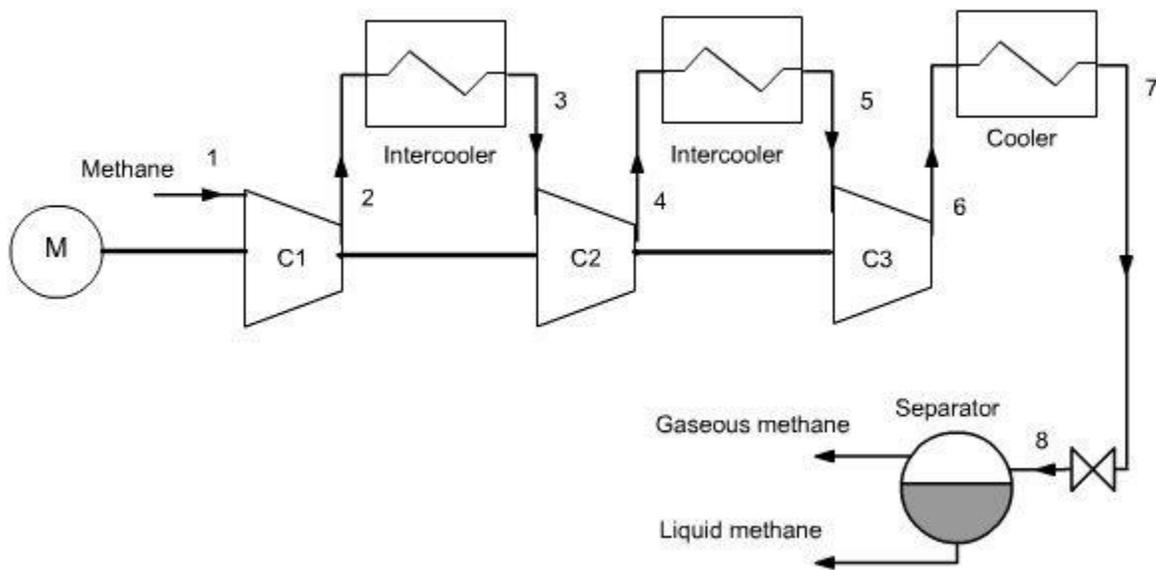


Figure 3 J-T Isenthalpic Expansion Cycle Example (Paris, s.f.)

2.3.2 Linde cycle

The Linde cycle (Figure below) improves the previous expansions by recycling the gas fraction in the separator and bringing it together with the methane compressed (before expansion) to a heat exchanger in such a way that the compressed gas needs not to be refrigerated down to -63°C by the cooler but to a higher temperature and in that economizing cooling power and optimizing the overall system efficiency.

In fact, by these adjustments, when running the calculation for this specific case, the work required per kilogram of liquified methane reduces about to the half than in the basic cycle.

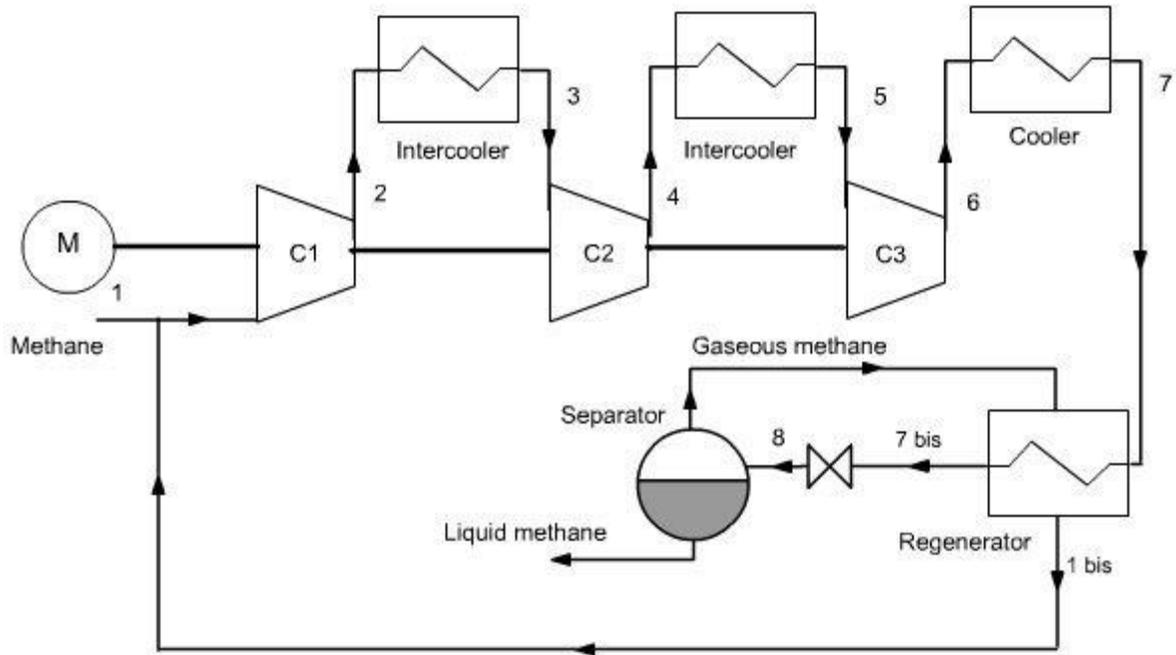


Figure 4 Lynde Cycle Modification to Example above (Paris, s.f.).

2.4 Reverse Brayton cycles

A reverse Brayton cycle achieves a cooling effect by reversing the gas turbine Brayton cycle: a gas is compressed, cooled, and then expanded as shown below (Figure 6). At the end of the expansion the temperature is low (J-T cooling principle).

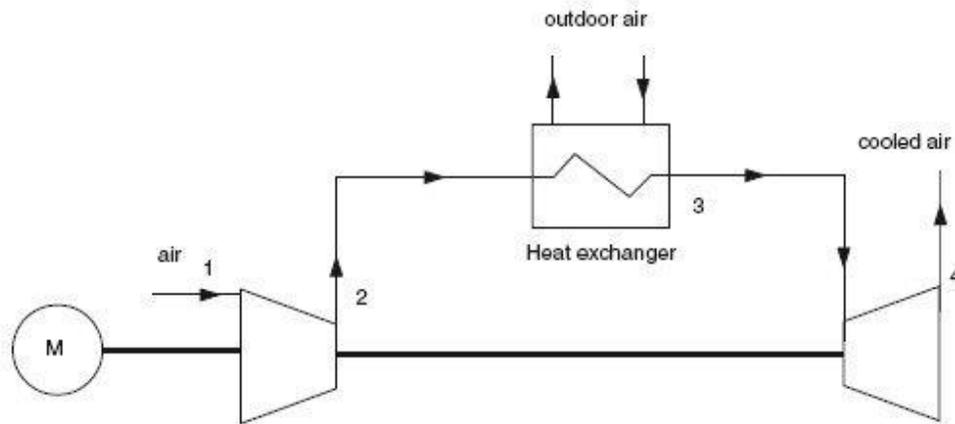


Figure 5 Reverse Brayton Standard Setup (Paris, s.f.).

In terms of thermodynamics the following T-S diagram describe what are the changes that the fluid undergoes throughout the Reverse Brayton cycle.

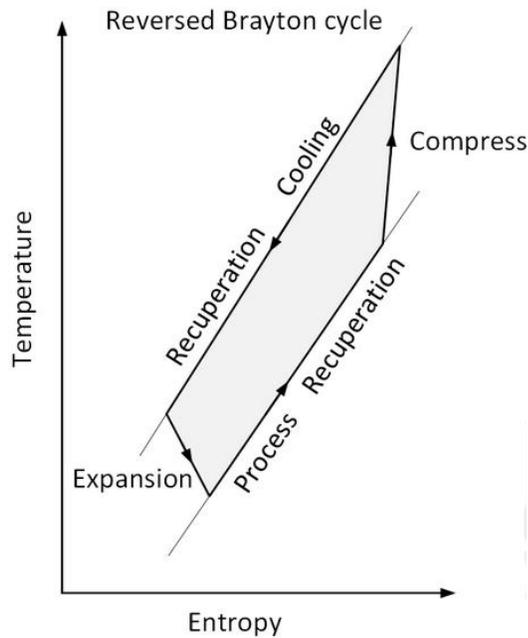


Figure 6 Thermodynamic Diagram of Reverse Brayton (Paris, s.f.).

An example was developed by B. Petit for producing cooling below $-253\text{ }^{\circ}\text{C}$ and the block diagram obtained is shown below (Figure 8).

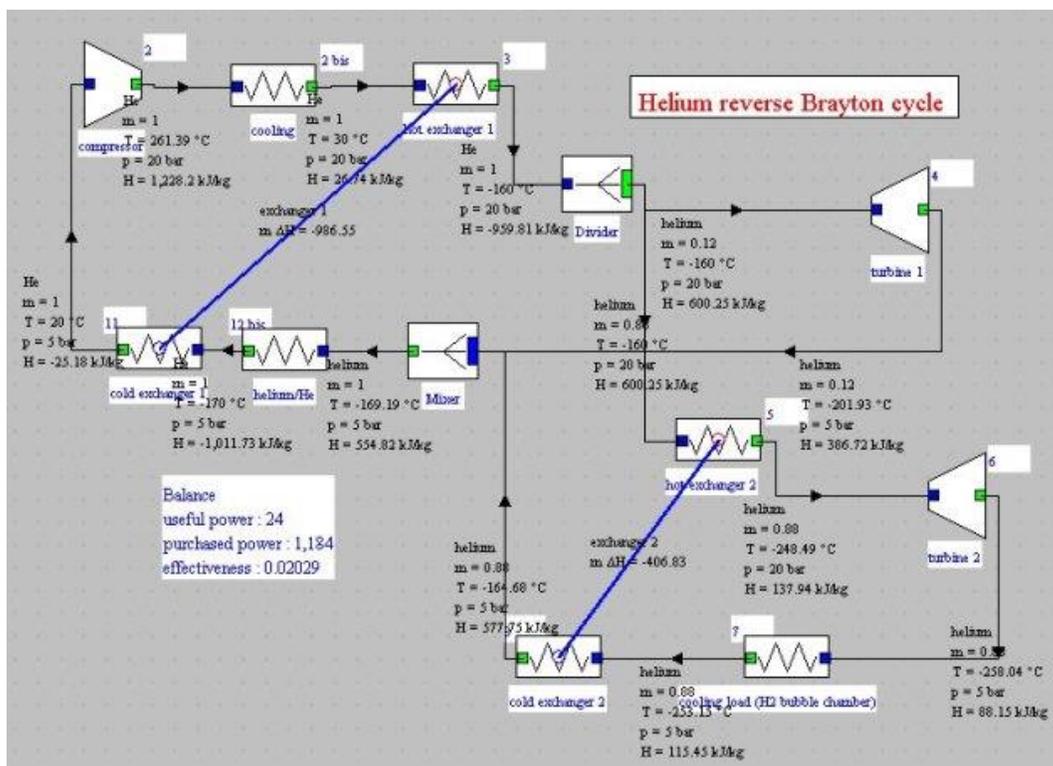


Figure 7 Reverse Brayton Cycle Example Diagram (Paris, s.f.).

In this cycle, helium is compressed to 20 bars, then cooled to 30°C before being divided into two streams which are expanded in parallel, the main stream following a conventional reverse Brayton cycle (bottom right of the diagram), while the secondary contributes to cooling the total flow.

In this case the cooled helium at -258°C is used to cool a load and the incoming fraction of the feed before entering the turbine for expansion (as before economizing cooling power and optimizing the overall system efficiency).

2.5 Mixed Processes: Claude cycle

Claude has proposed a cycle that involves a turbine and an expansion valve combining both isenthalpic and isentropic expansion in one model. This cycle has been used in many air liquefying facilities.

To illustrate this process, we use the work done by Santarelli in which hydrogen storage and liquification is modeled using Claude cycle.

As can be observed below (Figure 9), at its first stage the gas is compressed, and its temperature is raised. Thereafter is cooled and for this cooling work, nitrogen is used and is expanded through a turbine instead of a valve in a heat exchanger. This is because if it is assumed that the expansion is reversible and adiabatic, then that expansion in a turbine is isentropic, and the temperatures achieved by the process are much lower than those obtained by an isenthalpic expansion (valve). After cooled in this way the gas (hydrogen) passes through a valve in which undergoes an isenthalpic expansion (Joule-Thompson) producing the liquid.

At the latest stage of cooling the hydrogen is expanded through a valve at temperatures below its inversion temperature where undergoes cooling due to the Joule Thompson effect.

After this, once the liquified hydrogen is extracted, the gas fraction in the separator return to the compressor through the heat exchanger (Santarelli, 2015).

Below is a visual representation of the system.

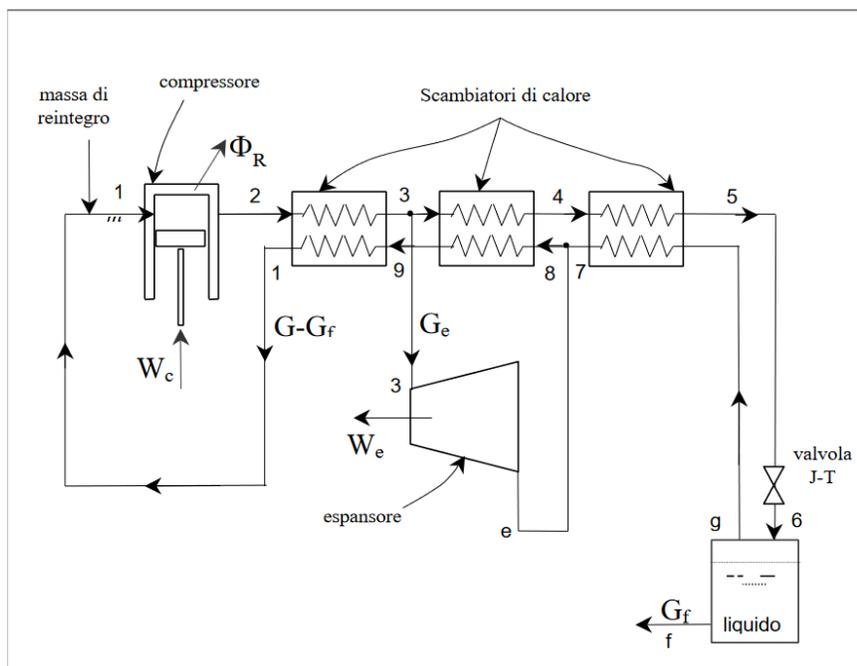


Figure 8 Claude Process Plant (Santarelli, 2015)

The advantage of this cycle is that the compression ratio can be significantly lower than in the case of the Linde cycle. One difficulty is that the expansion machine (turbine) cannot operate with good efficiency if the fluid remains in the vapor zone or keeps a high quality. The originality of the Claude cycle is to combine isentropic expansion in the turbine, and isenthalpic expansion only in expansion leading to the gas liquefaction.

Among the results found in this simulation highlights an optimum amount pressured inflicted by the compressor to the feed gas by which the maximum liquified fraction of the gas is found that in this case builds up to 52 bar corresponding to a 0.4682 liquified fraction of the feed gas (Santarelli, 2015).

2.6 Hydrogen production from a combined WGS and Gas Separator process

Another interesting article deals with the comparison between Case-I: Membrane based assisted and Case II: Cryogenic assisted, technologies employed for H₂ separation in a Hydrogen production from WGS process.

They analyze them in terms of the energy, exergy, and economic aspects of the processes. In both cases, H₂ was first produced from synthesis gas via water gas shift reaction and was then separated from other components using the respective mentioned methods.

Following is the diagram of the process.

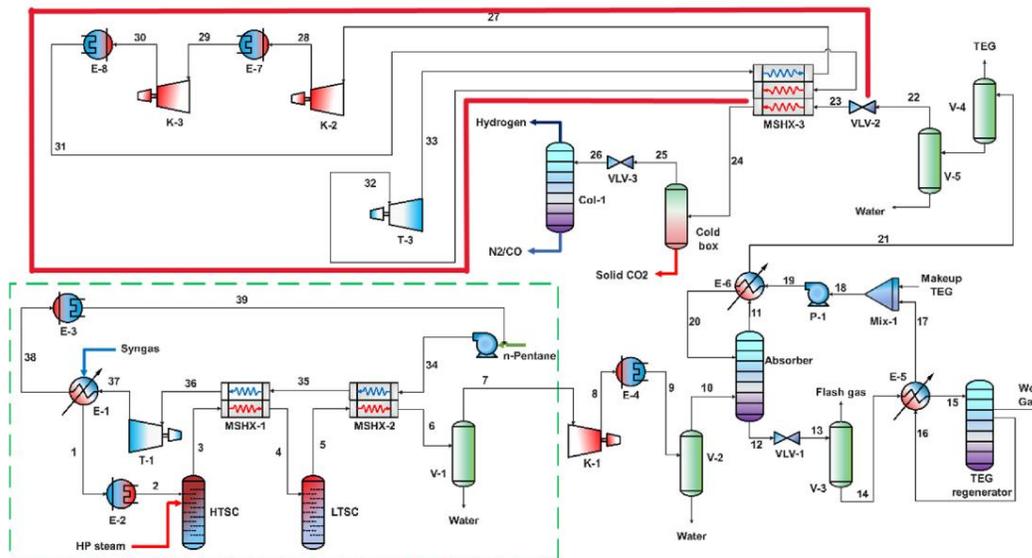


Figure 9 Diagram flow of Case-II (i.e., cryogenic process). (Ahmad Naquash, 2022)

The WGS (Water gas shift) reaction and ORC (Organic Rankine cycle) system sections are indicated by the green dashed box. Since the WGS reaction is exothermic the excess heat removed is used to generate power through the ORC.

The dry, H₂-rich SG stream (22) is then directed to the cryogenic unit. The stream temperature and pressure are first reduced to -58 °C and 4.5 bar, respectively, to convert CO₂ from a vapor to a solid, followed by the removal of the solid CO₂.

The SG stream (25) comprising CO, N₂, and H₂ is sent to the cryogenic distillation column to separate H₂ with a high purity and recovery. H₂ is extracted from the top, while CO and N₂ are separated from the bottom (liquid). The energy required to decrease the temperature of the SG stream for cryogenic processing is provided by the refrigeration cycle (outlined in red in Figure 4).

The software Aspen Hysys was employed to simulate both processes. Energy analysis reveals that case-I has a lower energy consumption (0.50 kWh/kg) than case-II (2.01 kWh/kg). However, low H₂ purity and recovery rates are the main limitation of case-I. In terms of exergy, the H₂ separation section in case-I exhibited a higher efficiency (28.4%) than case-II (14.7%). Furthermore, the economic evaluation showed that case-I was more expensive (\$17.7 M) than case-II (\$10.2 M) because of the high cost of the compressors required in the first.

Chapter 3 Methodology

As we already stated in the Goal and Scope section the thesis aims the modeling of a debinding system to be used at the PRI of the Italian gas network and feeding the model with the actual flows and pressures data obtained from the operator. This will serve as the input to the model for Italy use case and evaluate its technoeconomic feasibility.

Our specific modelling case will be a cryogenic debinding where the feed gas is composed of 80% hydrogen and 20% methane that undergoes a multistage cryogenic separation configuration.

The whole idea is to reach cryogenic temperatures at which the methane condensates and the hydrogen remains as gas in the separator vessel and in this way obtain hydrogen in gas phase with increasingly high purity until reaching temperatures down to -180°C (target) where hydrogen is recovered with $>98.5\%$ purity.

As it was seen in the Literature Review Chapter there are several mechanical cryocooler options that can meet the refrigeration requirements among which 3 highlights as the most important namely:

- a) isenthalpic expansion Joule-Thomson processes.
- b) isentropic expansion reverse Brayton cycles.
- c) mixed processes involving isenthalpic and isentropic expansion (Claude cycle).

In the Literature Review section, it was pointed out that although the reverse Brayton cycle produces a near-isentropic expansion by expanding the gas through a turbine (Fraser, 2006), sometimes the others may match better the specific characteristic of the process needed.

In fact, in the case described in this report it has realized that the J-T expansion worked better for this specific application even though initially it had started working with the classical reverse Brayton cycle to leverage from the higher temperature reduction (as a consequence of the near-isentropic expansion) and also the mechanical power produced when the gas expands in the turbine blades (as opposed to a JT expansion in which the gas can be thought as only doing work on itself and does not produce work) (Fraser, 2006).

As depicted below the cycle we are modeling typically consists of a low- and high-pressure buffer volume (LPBV and HPBV), a recuperative heat exchanger, the turboexpander (or a valve), and the load heat exchanger.

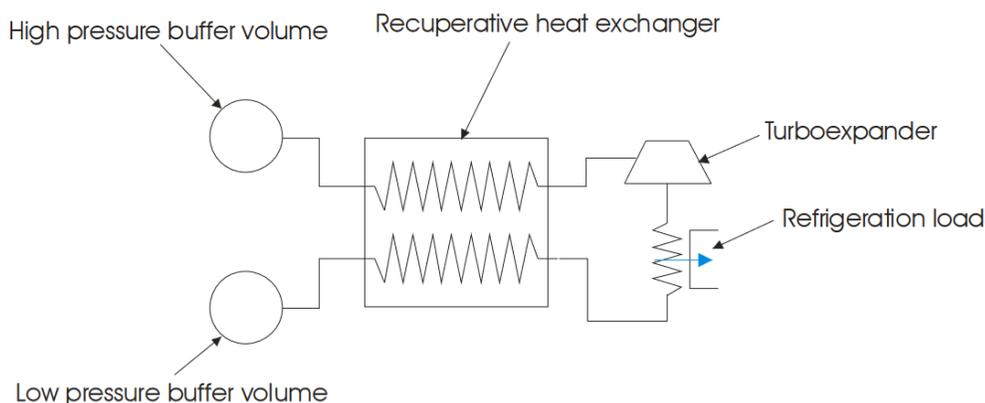


Figure 10 Reverse Bryton cycle schematic.

JT expansions, that refers to those expansions done using valves, can also reach temperatures as low as those reached by the turbines provided that the gas expanded is at temperatures at around -115°C meaning that they are already low. We acknowledged, performing the trials for this report setup (which will be explained in detail), that the use of valves had only a minor effect on the overall system performance.

Therefore, the JT cycle end up being useful for our application of reaching cool temperatures that causes gas separation in a mixture of hydrogen and methane (at these temperatures reached by the system the methane becomes liquified while hydrogen remains as gas making possible the separation).

Also, adapting the Figure 9 to the system which will be modeled it must be noted that the HPBV is the gas transported at high pressure upstream to a specific PRI or offtake in the NG grid, where pressure drops occur. This can be considered as a continuous flow that does not run out as it is provided by a central grid operator. The LPBV on the other hand is the gas that continues being transported downstream after a PRI or offtake point. Finally, compared to Figure 2 we consider no refrigeration load but rather the cooling of the fluid itself as the refrigeration load.

Since the high-pressure feed gas is deviated towards the deblending facility and the residual gas need to be delivered back to the mainstream in the grid, there is the need to deliver it back at a certain pressure (set by the grid operator) after the specific node where the separation is taking place. For this it becomes interesting to achieve that the residual gas that exits the cryogenic process the deblender facility is as much as possible only expanded down to that outlet pressure set by the operator so there is less need for recompressing, a feature that will be explained better in later sections.

Chapter 4 Use case modeling

Underlying the feasibility of reaching such low temperatures relies the benefit derived by constant recirculation of streams coming from different cooling stages downstream that return to be delivered back to the grid but not before transferring cooling energy to the fraction of the feed gas that is further being cooled in the process. That is, the cooling power of a cascade heat exchanger setup is higher than expanding the totality of the pressurized feed gas down to the lowest pressure in just one go. That is one of the reasons behind why the blending gas is cooled in a multistage fashion, to end up with a gas at high hydrogen purity.

Following is the setup we follow to model this system that is conversant with the modelling done in the GB work deblending feasibility analysis.

The basic principle behind this cryogenic gas separation is illustrated in the figure below.

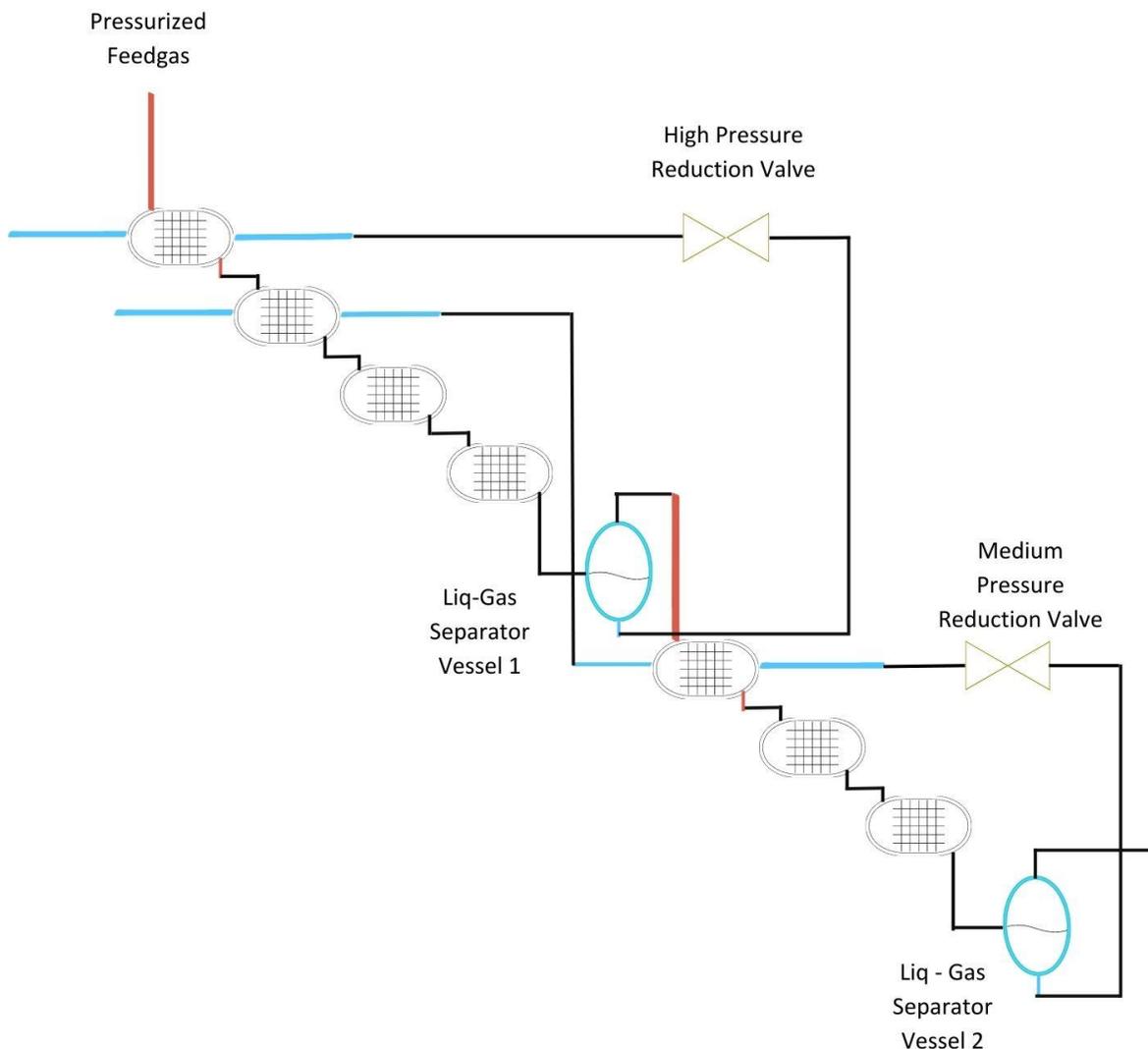


Figure 11 Basic operation principle, cryogenic gas separation

As it can be noted in each cooling stage, after going through the HX's where it gets cooled, the feed mixture is separated in a Separator vessel where the liquid fraction corresponding to the methane part is expanded in a valve and delivered upstream as a residual gas (it does not longer undergoes the cooling process). This cooled expanded methane serves as the cooling source for the J-T cycle and it gets evaporated as it goes through the heat exchangers to end up delivered back to the grid. The vapor fraction from the Separator continuous to the next cooling stage ultimately achieving the product separation (hydrogen).

In this way, the present thesis will allow us to take the Italian NG distribution reality and analyses its specifics debblending technical and economic feasibilities. For this we will take real data of the main flows and pressures at specific points of the Italian natural gas grid.

Important to note is that this process has already been simulated for the NGT of GB, which results have already been put out for public review, outputting an overall positive tecno-economic evaluation of this system.

4.1 Cooling Description Model Cases

In each cycle the fluid undergoes a multistage separation configuration, each one constituted by a set of heat exchangers, a separator vessel, a splitter, and an expansion valve. There is no unique way of putting together these components in each stage and the deciding factor which will decide among them is going to be how well the cooling is carried out, the number of cycles needed, and the amount of compression work required.

Therefore, before setting the chosen model setup as it is shown in Figure 11, three setups were considered from the beginning of the modelling and are explained briefly.

4.1.1 Case I

The feed gas enters the stage after being cooled through the previous stage heat exchangers and it is sent to the valve where it is expanded down to a certain pressure. After this the cooled gas enters the Separator Vessel where the liquid part (methane) is delivered upstream as cooling and the gaseous part enters a splitter where a fraction is taken to join the LNG and help cooling. The remaining gaseous part continues further down the cooling stages.

The drawback here is that less hydrogen is recovered, and the benefit is that higher fractions of the feedgas are kept at higher pressures.

The remaining gaseous part is also at ambient pressure at the last stage.

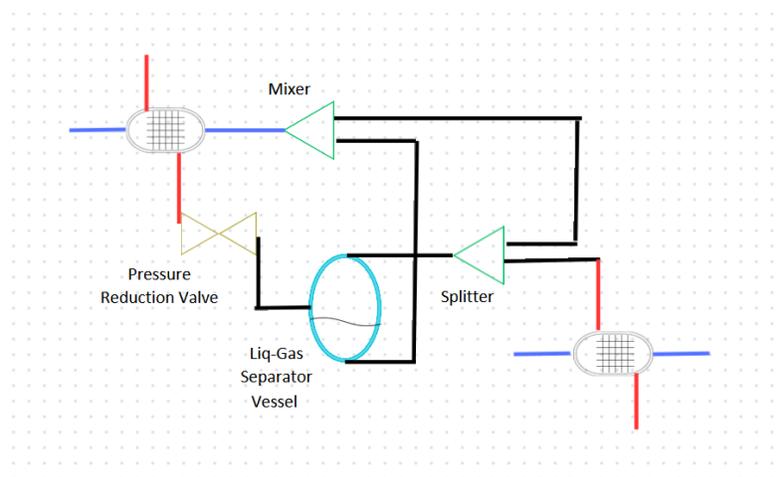


Figure 12 Model Case I

4.1.2 Case II

Same as Case I however, there is no imposition on recycling a fraction of the gas exiting the SV's and the recycled stream for cooling correspond only to the LNG exiting the SV's.

The drawback here is that the recycled feed gas fraction is lower at high pressures, however the benefit is that the HX's areas will be decreased and hydrogen recovery increases.

The remaining gaseous part is also at ambient pressure at the last stage.

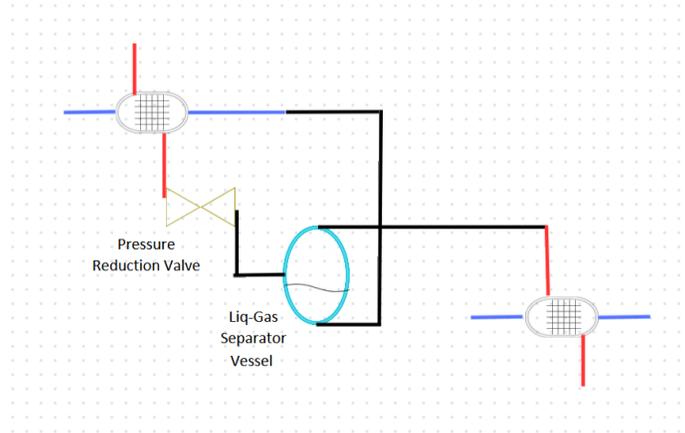


Figure 13 Model Case II

4.1.3 Case III

The feed gas enters the stage after being cooled through the previous stage heat exchangers and it is sent to the separator vessel where the liquid part (methane) is expanded through a valve down to a certain pressure and delivered back to the grid providing cooling all its way through. The remaining gaseous part in the separator vessel continues further down the cooling stages.

The remaining gaseous part at each stage conserves the initial pressure of the feed gas unless purposely the feed gas is brought down to achieve better efficiencies.

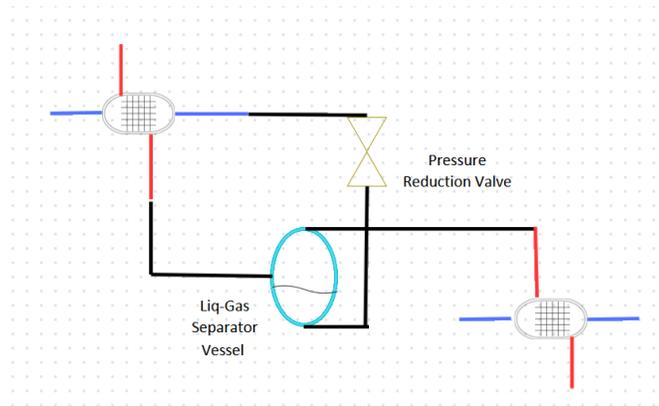


Figure 14 Model Case III

4.2 Similitudes among the models

What these three cryogenic separator models have in common is that in each cooling cycle setup the feed stream undergoes through the HX's setup to be cooled and then transferred into a separator vessel where the liquid fraction is withdrawn as a LNG and expanded to provide cooling and the remaining gas fraction continues to the following stages of the cooling process or/and a fraction of it is used to join the LNG stream.

In all these setups it is shared also the aim of maintaining the highest possible fraction of the residual streams (returning from all stages) at high pressure which has the benefit of reducing as much as possible the need to recompress them afterwards to meet the grid standards pressures required. And this is also one reason of the multistage cooling setup because all the residual gas streams returning from the stages have been expanded to low, medium, or high pressure facilitating the control and optimization over the fraction of the residual gas that we want to keep at higher pressures.

However, there is a cost to this that needs to be taken into account. If the residual gas coming from downstream has bigger fractions at the output pressure the system will demand more heat exchanger area (since they have been expanded less, they would be not so cold). Therefore, there will be requirement for more area in the HX for the stream to adequately exchange heat at the HX. However, again there is a benefit in this because this stream (from the first separator vessel) has just expanded down to the pressure required by the grid operator to be delivered after PRI and thus less compression power will be required.

Likewise, in the case we have bigger fraction of the residual gas expanded to lower pressures we have colder temperatures in these streams coming from the downstream cooling stages and therefore with less area in the HX's they would be able to transfer the same cooling energy to the feed gas and thus less HX equipment volume will be required.

That is, if we strive to retain a big mass flows fraction of the initial feed at pressure that meets the delivery pressure at that point in the grid, that fraction would not require recompression after.

Therefore, there is a tradeoff between the cost of recompressing and the cost of bigger heat exchangers equipment (which will be given a sense with a sensitivity analysis).

The residuals streams are brought together with the feed gas (or a fraction of it) into a set of heat exchangers where they cool the gas leveraging from these residual gas streams carrying cooling energy. These streams are then delivered back to the grid as a residual gas at almost near ambient temperature and at the pressure required for the grid at that point (weather this is accomplished because of the expansion being done down to the pressure required or with the use of recompression power as just explained).

Chapter 5 Selected Model Setup (50-5 bars) Description

The feature of the gaseous fraction that further undergoes the cryogenic cooling remaining at high pressure is what allows the last of the three models explained above to achieve the best performance.

Therefore, that model has been selected and is the one that will be explained in detail.

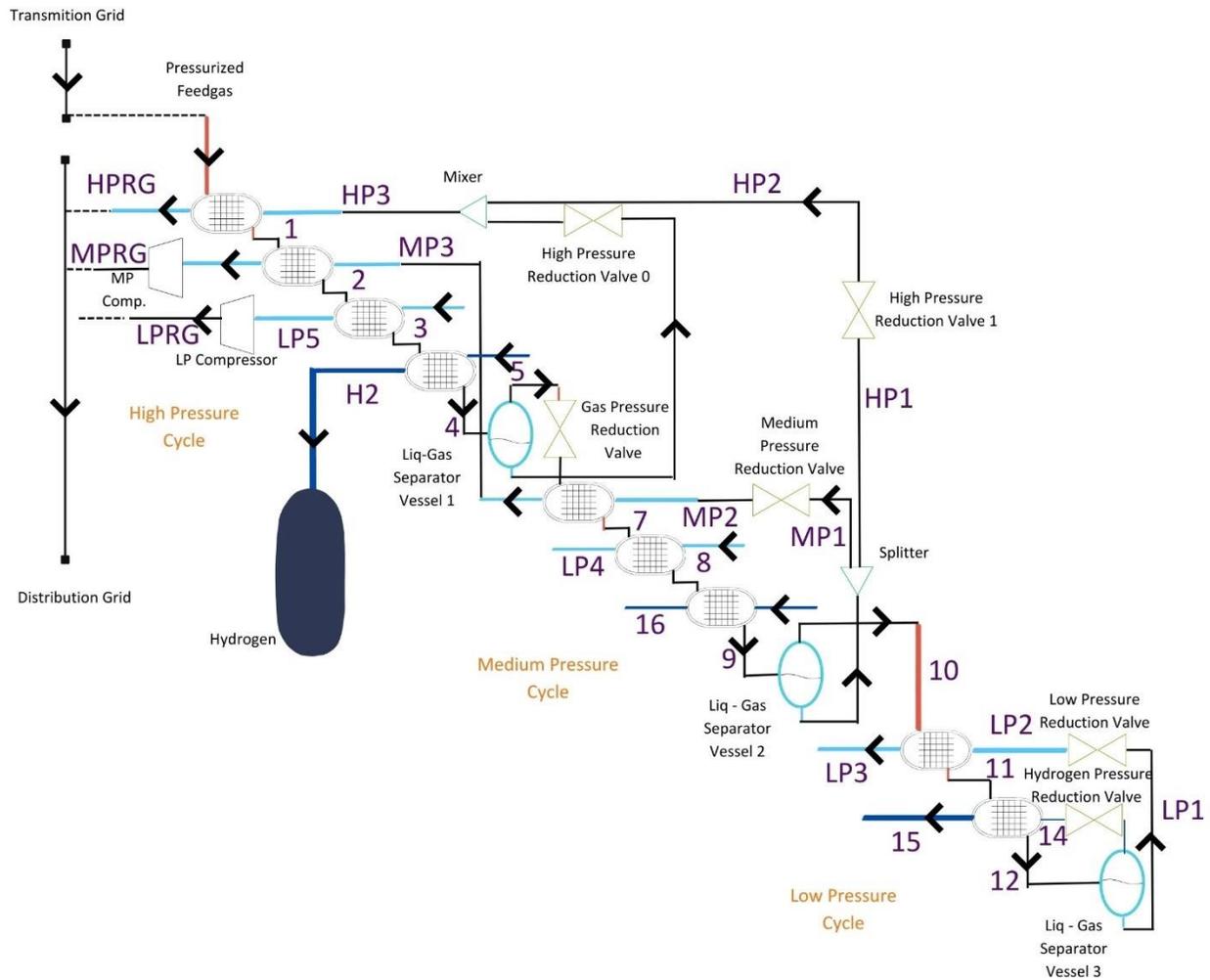


Figure 15 Cryogenic full cycle setup

A three-cycle stage cryogenic process is depicted above with each stage labeled as High, Medium and Low-Pressure Cycle.

To begin with it is considered a feed gas composed of 80% mol methane and 20% hydrogen that is at 50 bars (transmission grid) and need to be delivered next (distribution grid) at 5 bars which corresponds to representative inlet and outlet pressures at the PRI points that are managed in the gas distribution by the Grid operator in Italy. The mass flow corresponds to 1 MCMD (million cubic meter day) which is also representative in the gas distribution grid in Italy.

However, it is important to note that regardless of the fraction composition or initial pressures for the feed gas are considered the simulation can be replicated without extra troubles as it only requires changing the set of input values in the model.

In the system depicted in Figure 15, the pressure of the feed remains the same throughout the stages, the temperature of the feed gas after the first set of heat exchangers would decrease gradually.

First the feed gas enters the first set of heat exchangers with each one corresponding to the high, medium and low-pressure cooling since the cooling stream providing refrigeration in them correspond to the residual gas that has been expanded in that very stage or lower to the referred pressures (high, medium and low) respectively. Also, the last heat exchanger corresponds to the product cooling (H₂), since it is the stream from the last separation stage corresponding to the obtained product hydrogen.

In the Separator vessel 1 the liquid part is expanded down to a high pressure which will serve as the cooling stream for the first heat exchanger of the first set and which is generally attempted to set at the output pressure requested by the operator. In the other hand, the remaining gaseous part of the feed gas continues to the second stage but before doing it and only exceptionally the gaseous part is reduced in pressure arbitrary (in the Gas Pressure reduction valve) as this will permit having lower pressures in the lower stages Separator vessels that will facilitate the desired separation.

In the second stage the process follows the same logic as in the first one but only medium, low pressure and H₂ heat exchangers are in place since there are no expansions to high pressure in this stage nor in the following. The difference with the first one though is that once it has been cooled in the heat exchanger and after enters the separator vessel, the liquid part (methane) goes into a splitter in which a fraction only is expanded through a valve down to the medium pressure, but the remaining fraction is expanded through a valve down to the high pressure. Then these two streams are delivered upstream as cooling energy as usual and while the stream that was expanded to medium pressure will serve as the cooling stream in second heat exchanger of the first set, the higher-pressure stream is brought to the first heat exchanger of the first set with the high-pressure residual gas stream from the upper stage (that was also expanded down to that pressure). This extra creative block management is done to retain a greater fraction of the total residual gas at higher pressure and in that way reduce recompression power as was already explained in more detail in the previous section.

Lastly, the remaining gaseous part continues to the third stage where passes through low pressure and H₂ heat exchanger (only low-pressure expansion in this stage) and reaches the separator vessel at the desired temperature of -175°C in which the hydrogen (gas part) is almost entirely separated from the methane (liquid part). The latter is expanded down to low pressure and is in turn delivered back to the grid passing through all the LP heat exchanger upstream (third, second and first heat exchanger for the three cycles respectively) and finally entering a compressor to be compressed at the high pressure that is the delivery pressure at that point set by the grid operator. Likewise, the H₂ stream is expanded down to 20 atm (standard arbitrary chosen product pressure) to then gain heat by passing through all the H₂ heat exchangers in upper stages to be obtained at near ambient temperature to be used for the desired application.

5.1 Thermodynamic of the process

5.1.1 Pressure – Enthalpy analysis (P-H)

In terms of thermodynamics the expansions from 60 bar to 20 bar in the P-H diagram for methane are seen. We can see that for initial temperatures of -120°C shows how the use of a turbine or a valve becomes remarkably irrelevant as there is not much enthalpy difference between the line that follows an

isentropic/turbine expansion (red line) and the line follows an irreversible/valve expansion (green line). On the contrary as we move toward the right of the bell the difference in enthalpy between these expansions start to become increasingly significant as it can be noted when we see the red line (corresponding to isentropic expansions) reaching lower temperatures. This justifies the use of valves in our model as our expansions are located in the left region corresponding to cryogenic temperatures from -85°C down to -175°C and therefore no much further cooling is achieved by the use of a turbine compared to a valve.

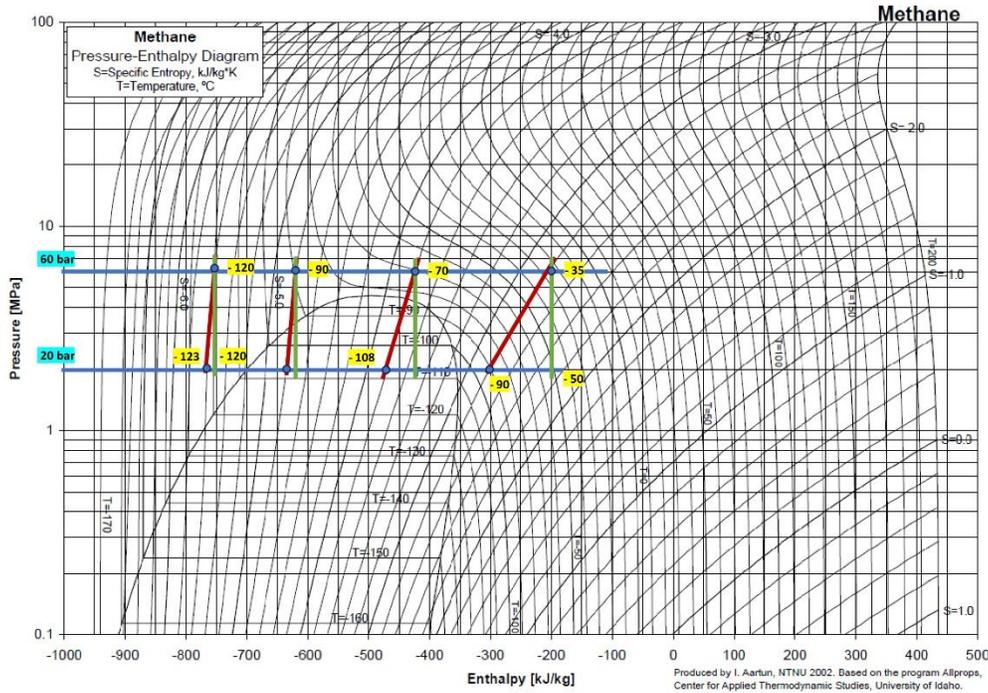


Figure 16 Thermodynamic P-H Diagram of Cryogenic cycle expansion

5.1.2 Enthalpy – Entropy analysis (H-S)

We can again observe the expansion cycle from the thermodynamic point of view but now seen on the Enthalpy-S diagram. Basically, we can observe a similar pattern than in the P-S diagram. In an expansion from 60 to 20 bar the enthalpy difference that results from an isentropic expansion (red line) and an irreversible expansion (green line) and the temperatures reached are not so significant for lower temperatures. That is backed up by observing the difference in the entropy axis where it can be observed that the work destruction is not so significant compared the one that results form an expansion at a temperature of for example -35°C that can be observed in the diagram. This again justifies the use of valves in the model rather than turbines.

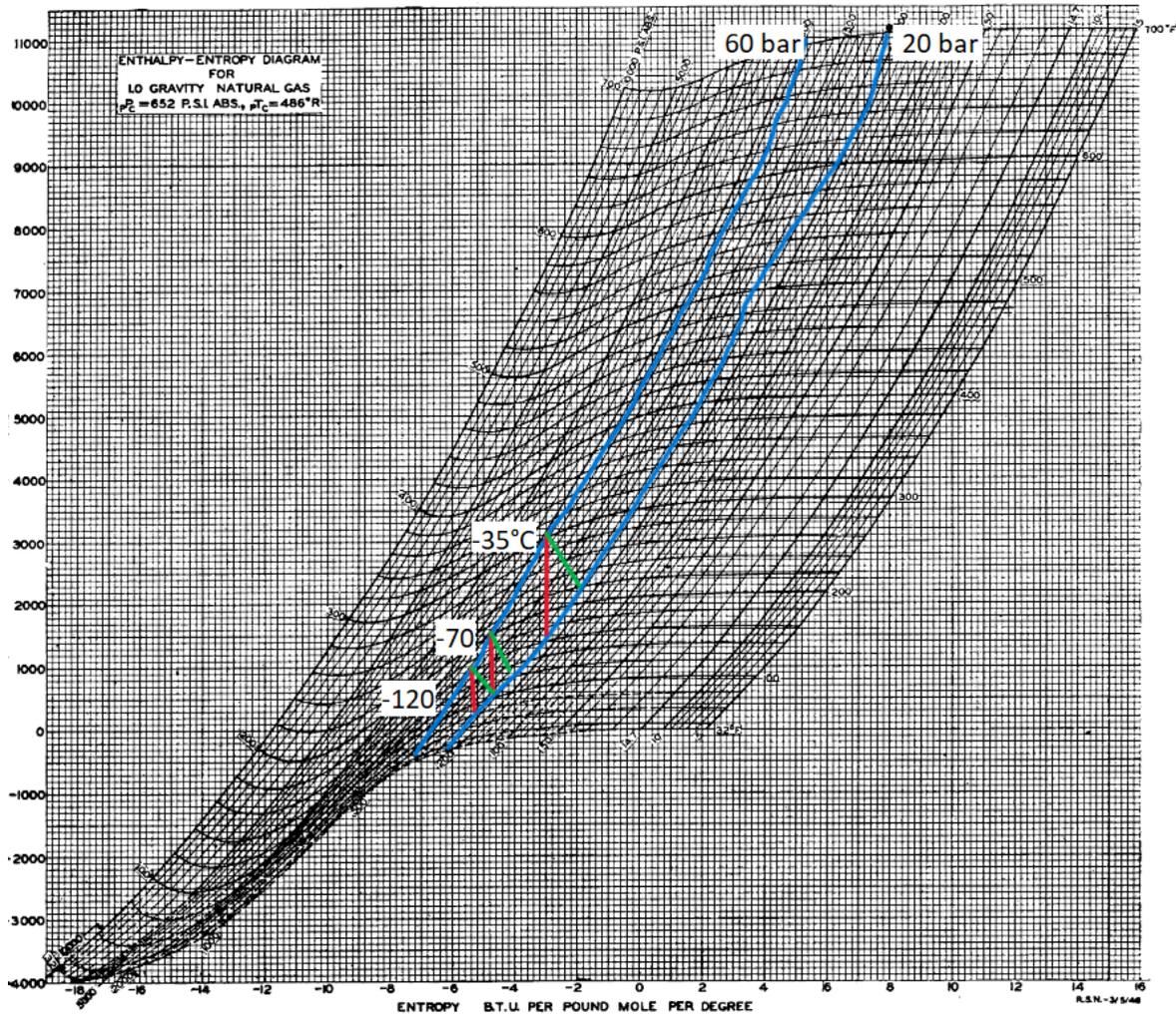


Figure 17 Thermodynamic H-S Diagram of Cryogenic cycle expansion

5.1.3 Inversion temperature consideration

Hydrogen has a boiling point of $-252^\circ C$ and methane has one of $-162^\circ C$. Depending on the gas composition to be expanded the effect of the expansion may be one of cooling or heating due to the Joule Thompson effect. For all the chemical species except helium and hydrogen the pressurized fluids undergo cooling when expanding at around ambient temperature and pressure. In the case of hydrogen, however, it gets heated at these conditions but regardless for any real gas expanded adiabatically still exists an inversion temperature below which the fluid gets cooled. For hydrogen this temperature is $-73^\circ C$ and is denominated as its inversion temperature. In our case, however, due to our configuration of HX's in which we take advantage of fractions of the initial feed gas that become residuals streams coming cooled from the intermediate cooling stages, we are able to reach the first turboexpander with the feed gas at temperature that is below the inversion temperature. Therefore, the main feed gas when expanded through the turbine goes down in temperature, both the methane fraction (which also does it at ambient temperature and pressure) as well as the hydrogen fraction for the reason just outlined.

Chapter 6 Model Use Case (50-5 bars) Results

As we stated before we are modeling the case where the feed gas is at 50 bar and the residual gases resulting from each stage need to be delivered back at 5 bars. This is following the actual pressures the Italian grid operator manages.

The internal pressures chosen at each valve to work the cryogenic process are shown in the table below:

Table 2 Model Internal chosen pressures drops.

Valve							
Name	GASPRV	HPPRV0	HPRV1	MPRV	LPRV	HPRV	
Defined inlet pressure [bar]	50	50	30	30	30	30	
Defined outlet pressure [bar]	30	5	5	5	1.5	20	
Calculated pressure drop [bar]	20	45	25	25	28.5	10	

The result of the model below outlines the hydrogen produced and relevant equipment metrics related to it:

Table 3 Product (Hydrogen) Results main metrics

Hydrogen recovery	90%
Hydrogen molar flow (kmol/hr)	308.85
Hydrogen mass flow (kg/hr)	617.7
Purity Hydrogen (%)	98,11
Comp Work (kW)	329.4
Total HX' area (m2)	343

For this specific case the medium pressure reduction valve outlet pressure is set at 5 bar just at with the HP pressure reduction valve as the benefits were minor compared to the significant advantages whenever no compressor power at the medium pressure residual gas is required (outlet pressure required by the grid operator is 5 bar).

Now as we follow the fraction of the feed gas that undergoes the cooling until the last separator vessel, it can be seen how the temperature continuously decreases until it reaches -173°C at the entrance of the last separator vessel (Table 1, stream 12). After that it returns as a refined H2 through the streams from 13 to 16 and H2 gaining temperature as it combines in the H2 heat exchangers.

Table 4 Gas Temperature cooling evolution

Stream	T(°C)	Stream	T(°C)	Stream	T(°C)
FEEDGAS	15	6	-96,157	12	-172,936
1	-57,435	7	-108,847	13	-174
2	-65,865	8	-109,048	14	-175,647
3	-74,855	9	-110,026	15	-173,427
4	-78,476	10	-115	16	-110,048
5	-81	11	-172,427	H2	-75,855

Note also that the temperature of the product hydrogen ready for use is still cold (-75°C) and external heating might be considered if necessary for the application.

Moreover, we can observe the molar fraction in those enumerated streams and see how the hydrogen molar fraction goes increasing correspondingly from the beforehand set 20% until the stream 13 reaches 98,12% H₂ mol (as the temperature decreases the methane condenses and is withdrawn from the mixture as liquid in the separator vessels). In this case the molar fraction remains naturally constant from stream 13 to 16 and H₂ (no separator vessels in this section of streams).

Table 5 Gas Hydrogen Molar Fraction Evolution

Stream	H2 mol%	Stream	H2 mol%	Stream	H2 mol%
FEEDGAS	20	6	20	12	39,6098
1	20	7	20	13	98,1153
2	20	8	20	14	98,1153
3	20	9	20	15	98,1153
4	20	10	39,6098	16	98,1153
5	20	11	39,6098	H2	98,1153

Below in Table 5, in an analogic fashion it can be seen the residual gases exiting the separator vessels at each stage, namely low, medium and high-pressure residual gases. First the residual gas is expanded which explains the decrease in temperature from the first row to the second but then we see how their temperatures go on increasing as they return to the grid passing through the heat exchangers cooling the feed gas.

Table 6 Residual Gases Temperature evolution

LP Res. Streams	T(°C)	MP Res. Streams	T(°C)	HP Res. Streams	T(°C)
LP1	-174	MP1	-115	HP1	-115
LP2	-173,888	MP2	-139,831	HP2	-139,831
LP3	-117,44	MP3	-99,582	HP3	-139,831
LP4	-109,847	MPRG	-58,518	HPRG	14
LPRG	4,691				

It can be noted here that the temperature of the final residual gas for low, medium and high pressure residual gas and we see that they reach similar temperatures than the grid temperature with the exception of MPRG which is at -58°C.

Likewise in Table 6, it can be seen how the molar fraction for these streams is constant CH₄ mol 97,7%. This is because they correspond to the fraction at the separator vessels that have condensed (Methane). Basically, they are LNG with relative high purity.

Table 7 Residual gases molar fractions

LP Streams	CH4 mol%	MP Streams	CH4 mol%	HP Streams	CH4 mol%
LP1	97,3538	MP1	97,7547	HP1	97,7547
LP2	97,3538	MP2	97,7547	HP2	97,7547
LP3	97,3538	MP3	97,7547	HP3	97,7547
LP4	97,3538	MPRG	97,7547	HPRG	97,7547
LPRG	97,3538				

6.1 Model Results Sensitivity Analysis

Following two variations in the discharge pressure at the stages are carried out from the base solution presented above.

6.1.1 Base Case

The base case is described below:

Table 8 Use Case Key parameters indicators for sensitivity analysis

Hydrogen mass flow (kg/hr)	617.7
Purity Hydrogen (%)	98,11
Comp Work (kW)	329.4
Total HX' area (m2)	343

Here the low-pressure cycle discharge pressure is 1,5 atm, the total HX's area is 343 m2 and the compression work required is 329.4 KW. From this case the pressure drops at the low-pressure cycle valve are changed to observe its effects.

6.1.2 Variation cases (cooling target fixed)

Variation in discharge pressure at LP cycle is described below:

Table 9 1st Variation Case Key parameters indicators for sensitivity analysis

LP Cycle	Pressure 1	Pressure 2	Pressure 3	Pressure 4
Pressure (atm)	1.5	1.1	0.6	0.2
Total HX' area (m2)	343	329	293	258
Comp Work (KW)	329.4	433.123	648.759	1155.29

Here the discharge pressure is decreased from 1,5 to 0,2 atm at intervals (the cooling target is fixed). Consequently, the LP cycle HX area goes on decreasing from 165 to 80 m2 bringing about a reduction in the total HX area from 343 to 258 m2; and naturally the compression work goes on increased from 329.4 to 1155.29 KW.

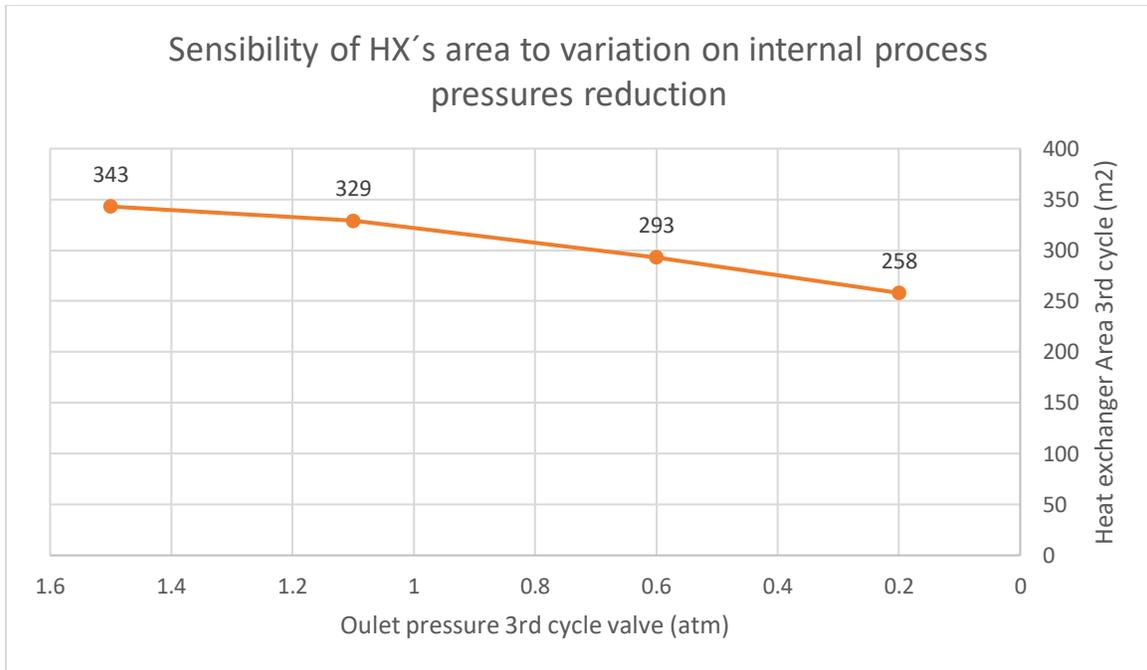


Chart 1 Sensibility of HX's area to variation on internal process pressures reduction

In Chart 1 it is observed that varying the discharge pressure to 0,2 atm means that out of the total residual gas flow part of it is being returned to the grid at lower pressures than before. In doing so the temperatures go further down and because of it in order to cool the feed gas at the heat exchanger less area is required, being beneficial in terms of the cost (less heat exchanger area).

However, in Chart 2 it is graphically visible that since the gas has been expanded further down in pressure the compression work required to bring it back to 5 (the operative delivering pressure) is higher which below.

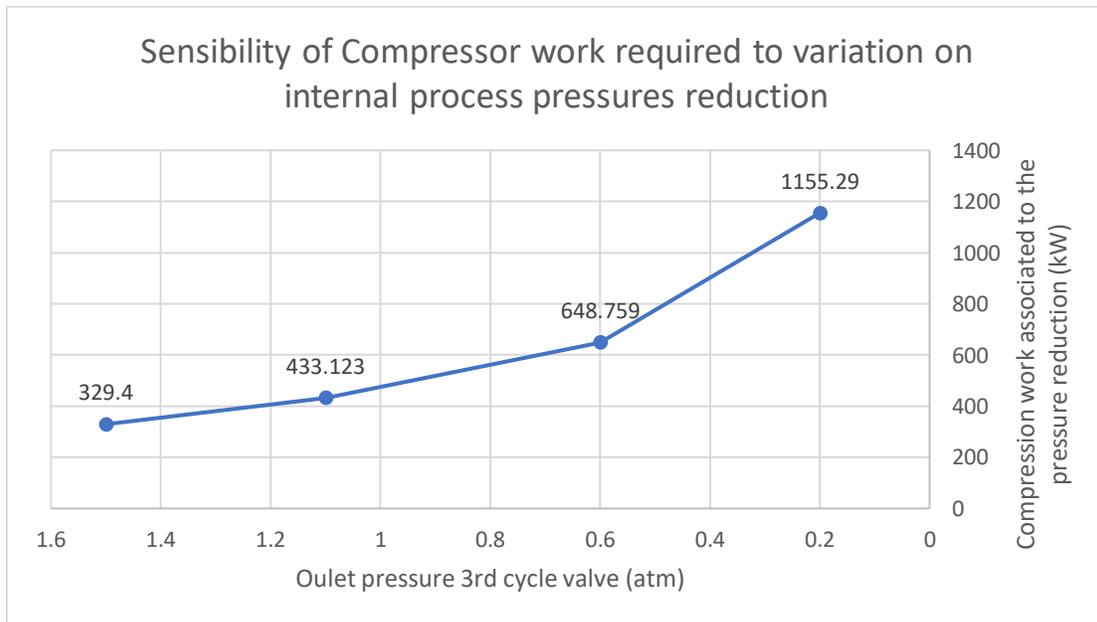


Chart 2 Sensibility of Compressor work required to variation on internal process pressures reduction.

6.1.3 2nd variation case (the cooling target is let free to vary)

Another relation that can be observed is between the variation in discharge pressure at LP but now seeing the effect on the temperature achieved at the 3rd cycle and consequently on the purity of the product stream:

Table 10 2nd variation Case Key parameters indicators for sensitivity analysis

LP Cycle	Pressure 1	Pressure 2	Pressure 3	Pressure 4
Pressure (atm)	1.5	1.1	0.6	0.2
Temperature (°C)	-174	-178	-181	-183
Purity Hydrogen %	98.11	98.72	99.05	99.24
Hydrogen recovery %	90.2	90.28	90.30	90.31

Here the LP discharge pressure is decreased from 1.5 to 0.2 again. Consequently, the temperature achieved at the last LP separator vessel goes on decreasing from -174 to 183 °C (Chart 3) and the purity of the gaseous fraction exiting the separator vessel goes on increasing from 98.11% to 99.24%.

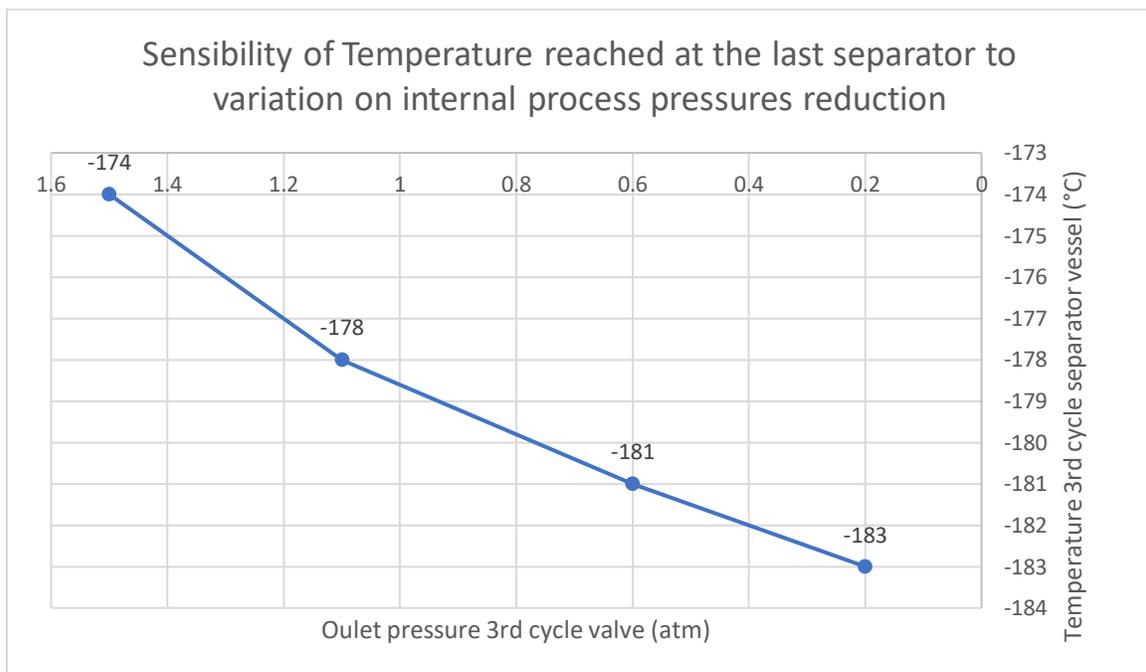


Chart 3 Sensibility of Temperature reached at the last separator to variation on internal process pressures reduction

Correspondently, in Chart 4 it is graphically visible that reaching lower temperatures has a positive impact in terms of higher product recovery and purity obtained.

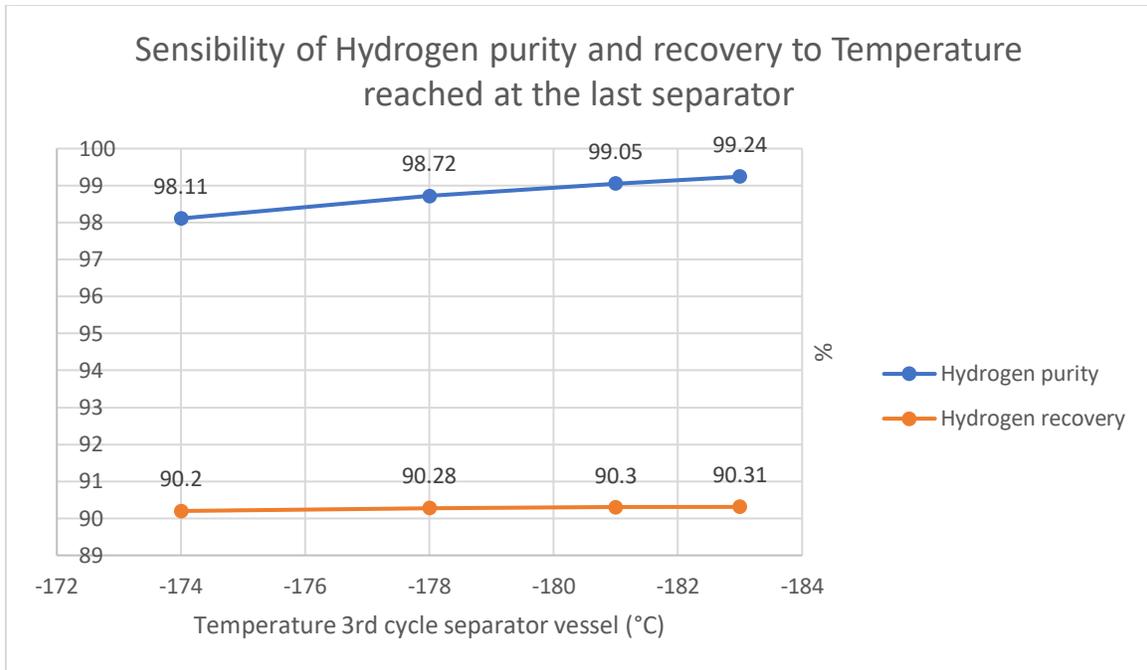


Chart 4 Sensibility of Hydrogen purity and recovery to Temperature reached at the last separator.

In conclusion, in one hand the first sensitivity analysis confirms this model's theoretical presumption of the existence of a tradeoff between the compressor power required and the size of the heat exchangers. As we move downwards in outlet pressures, we increase costs in compressors but decrease costs in Heat exchanger equipment.

On the other hand, the second sensitivity analysis establishes the tradeoff between the compressor power required and the rate of recovery and purity of the product. As we move downwards in outlet pressures, we increase costs in compressors but also increase the hydrogen recovery and purity.

Chapter 7 Technoeconomic Evaluation Use Case

In this section we are going to evaluate the costs of the case outlined in the previous section. Making use of the current market prices for the HX and compressors, we attempt to find the more economically convenient case.

7.1 HX's

The different types of Heat Exchanger are:

- a) Finned tube Heat Exchanger or Air-cooled Heat Exchanger. Suitable for: air/gas to fluid.
- b) Shell And Tube Heat Exchanger. Suitable for: fluid to fluid / fluid to gas.
- c) Plate Heat Exchanger or Gasket Plate Heat Exchanger. Suitable for: fluid to fluid / fluid to vapor.

(Sterlingtt, s.f.)

The most frequently used materials for the plates are stainless steel (AISI 304, 316), titanium and aluminum. (ONDA, s.f.)

Titanium has excellent corrosion resistance against seawater. Therefore, it is frequently applied to the primary members, such as heat-exchanging plates and piping, of plate-type heat exchangers (PHEs) that use seawater for cooling and heating. (Kobelco, s.f.)

Methane is non-corrosive and may be contained at ambient temperatures by most common metals used in installations designed to have sufficient strength for the working pressures involved. (Afrox, s.f.)

7.1.1 HX's Prices

Table 11 HX's Prices

Plate surface (m ²)	Price each (US\$)
120	36,000
140	39,000
160	42,000
200	49,000
220	54,000
260	61,000
280	65,000
300	71,000
350	80,000

(DACE, s.f.)

7.2 Compressors

The cost of a compressor is roughly: U.S. \$0.5–0.9 million/MW for the 5–15 MW range; \$0.8–1.5 million/MW for the 1–5 MW range; and \$1.3–4/W for units below 1 MW.

Historical data allow developing models capable of estimating installation cost within certain limits:

$$\text{Compressor Unit Cost} = A \times (\text{Compressor Package Cost}) + B \quad \text{Equation 1}$$

Where,

Factor A: Auxiliaries and accessories required for each compressor package (foundation, civil works, piping, additional steel structures, etc.); its value usually falls between 1.25 and 2.5.

Compressor Package Cost: Mounted facilities, such as the driver and lubrication oil system.

Factor B: Auxiliaries and accessories required for each compressor unit, such as protection systems, electrical facilities, safety equipment, unit steel structures, etc. (ALMAS, 2014).

7.3 Base Case Technoeconomic analysis

For this case the HX area required is 343 m² and by looking at the table above (Table 9), the cost associated to HX's is:

- 343 m²: \$78,600 € (around €80,000/350m²)

For this case the compressor power required is 329.4 KW and by looking at an option for a small compressor installation, the cost associated to the Compressor is:

- 330 kW: \$510,000 € (around €1.545/W)

Using equation 1:

As it is a small package, we estimate A to be 1.31 and B to be around \$97,270 €, the compressor unit cost is:

$$\text{Compressor Unit Cost} = 1.31 \times (510,000) + 97,270 = \$765,370 \text{ €}$$

Note that this cost comprises the engineering and installation for which they will need to be excluded when calculating the total cost of the deblender facility.

7.3.1 Capex

Table 12 Capex Breakdown

Deblender Facility Costs	Cost
CAPEX - Design & Engineering	214,558 €
CAPEX - Installation	286,078 €
CAPEX - Balance of Plant	238,398 €
CAPEX - Pipes and valves	126,595 €
CAPEX - HX's	78,600 €
CAPEX - Compressor	765,370 €
TOTAL	1,709,599 €

Assumptions:

- The Pipes and valves cost is 15 % of the Capex HX's and compressor.
- The Design & Engineering cost is 30 % of the Capex HX's, pipes, and compressor.³
- The Installation cost is 40 % of the Capex HX's, pipes, and compressor.⁴
- The Balance of Plant cost is 33 % of the Capex HX's, pipes, and compressor.⁵

7.3.2 Opex

Table 13 Opex Breakdown

Deblender Facility Operative Costs	Cost
OPEX – Electricity Consumption	360,693 €
OPEX – Labor	28,178 €
OPEX - Tax & insurance	51,287 €
OPEX – Depreciation	68,383 €
TOTAL	508,542 €

Assumptions:

- The facility is operating 15hr/day all the year.
- The cost for electricity in Italy for industry customer is 200 €/MWh (ARERA, 2023).
- Labor cost is 38.6 €/hr and an operator is required for two hours a day.
- Tax & insurance is considered to be 3% of the Total Capex.
- Depreciation is calculated considering a life span of the deblender of 25 years.

7.3.3 LCOH

The levelized cost of hydrogen (LCOH) is a measure of the average net present cost of electricity generation over the asset's lifetime.

$$LCOH = \frac{\alpha \cdot CAPEX + OPEX + \beta \cdot ABEX}{AHP} \quad \text{Equation 2 LCOE Formula}$$

Where,

- CAPEX: Upfront Capital Cost
- OPEX: Ongoing Operational Cost
- ABEX: End of life Cost
- AHP: Annual Hydrogen Production

The factors α and β account for the impact of timing on the CAPEX and ABEX costs compared with the life of the project and are defined as follows:

$$\alpha = \frac{r \cdot (1+r)^{Nop}}{(1+r)^{Nop} - 1}$$

³ Compressor Capex including only the compressor equipment itself.

⁴ Compressor Capex including only the compressor equipment itself.

⁵ Compressor Capex including only the compressor equipment itself.

- $\beta = \frac{1}{1+r} \cdot \frac{r}{(1+r)^{Nop}-1}$

Here,

- Nop : Operating life (yr)
- r : Rate of return (%)

Let consider that the rate of return desired from the investment is of 5% and that the deblending machine will operate for 25 years.

From these assumptions the financing parameters are:

- $\alpha = \frac{0.05 \cdot (1+0.05)^{25}}{(1+0.05)^{25}-1} = 7.1\%$
- $\beta = \frac{1}{1+0.05} \cdot \frac{0.05}{(1+0.05)^{25}-1} = 2\%$

Besides:

- The End-of-life cost ABEX⁶ is 15 % of the Capex HX's, pipes, and compressor.
- 617.7 kg /hr per hour are produced by the facility.

Therefore, the levelized cost of mass hydrogen produced by the deblending facility for our Base Case is:

$$LCOH = \frac{0.071 \cdot 1,681,963 \text{ €} + 506,607 \text{ €} + 0.02 \cdot 145,584 \text{ €}}{3,381,907 \text{ kg}} = 0.187 \text{ €/kg}$$

Finally, if used as a fuel for combustion, 1 kg of hydrogen releases 125,806kJ (McLean, 2022).

Then, the levelized cost of energy produced by the deblending facility for our Base case (when using hydrogen as a fuel for combustion) is:

$$LCOE = \frac{0.071 \cdot 1,709,599 \text{ €} + 508,541 \text{ €} + 0.02 \cdot 145,584 \text{ €}}{425,670,891 \text{ MJ}} = 1.486 \text{ €/GJ}$$

There is the need to point out that these levelized costs of energy are just the costs for deblending it giving for granted that the hydrogen is already there in this case in a mixture with methane. However, there has been a cost for obtaining the hydrogen which need to be added to this value.

Nowadays the kg of green hydrogen can be found at 10 to 15 € per kg, whereas grey hydrogen can be found at around 2 € per kg. However, it is estimated that with the massive deployment of renewables and its integration to the production of hydrogen, the price of green hydrogen will be available at around 2.5 €/kg and therefore the total cost to get the hydrogen thorough this process of deblending will be at around 3 €/kg.

7.4 Technoeconomic Results Sensitivity Analysis

7.4.1 1st variation case

The following variation HX's areas are considered which corresponding cost are calculated using the table above (Table 9):

- 329 m²: \$76,800 € (around 71,000/300m²)

⁶ Compressor Capex including only the compressor unit cost.

- 293 m²: \$67,600 € (around 65,000/280m²)
- 258 m²: \$60,600 € (around 57,000/240m²)

The following variation compressor powers are required which corresponding costs are calculated above:

- 433.123 KW: \$616,117 € (around € 1.4225/W)
- 648.7 KW: \$843,310 € (around € 1.3/W).
- 1150 KW: \$1.035 million € (around € 0.9/W).

Following the procedure as it was done for the Base Case which values are already known we come up with the table below summarizing the LCE for each case along with other previously known parameters:

Table 14 Information for Technoeconomic Sensitivity Analysis

	Pressure 1	Pressure 2	Pressure 3	Pressure 4
Pressure (atm)	1.5	1.1	0.6	0.2
CAPEX (€)	1,709,599	2,029,615	2,649,001	3,753,434
OPEX (€)	508,541	644,520	923,934	1,550,168
Hydrogen recovered (kg/hr)	617.7	617.87	618	618.1
LCO Mass H2 (€/kg)	0.187	0.234	0.329	0.538
Comp Work (KW)	329.4	433.123	648.759	1150.29
Hydrogen purity	98.11	98.72	99.05	99.24

Here the LP discharge pressure is decreased from 1.5 to 0.2 atm. Consequently, the hydrogen recovery increases from 617.7 to 618.1 kg/hr. Varying the hydrogen recovery means naturally that this will tend to decrease the LCOH, however this is achieved by means of decreasing the discharge pressure at the LP cycle which will increase the compressor work required afterwards and which in turn will tend to increase the LCOH. Additionally, the purity of the gaseous fraction exiting the separator vessel goes on increasing from 98.11% to 99.24% consequently.

We can see in Chart 6 how the compressor work outweighs the hydrogen recovery in determining the LCOH. Besides, it does so in an exponential way meaning that the benefits of getting more pure hydrogen are extremely costly and becomes increasingly higher. This suggests the option of having a separate purifier after this deblender setup.

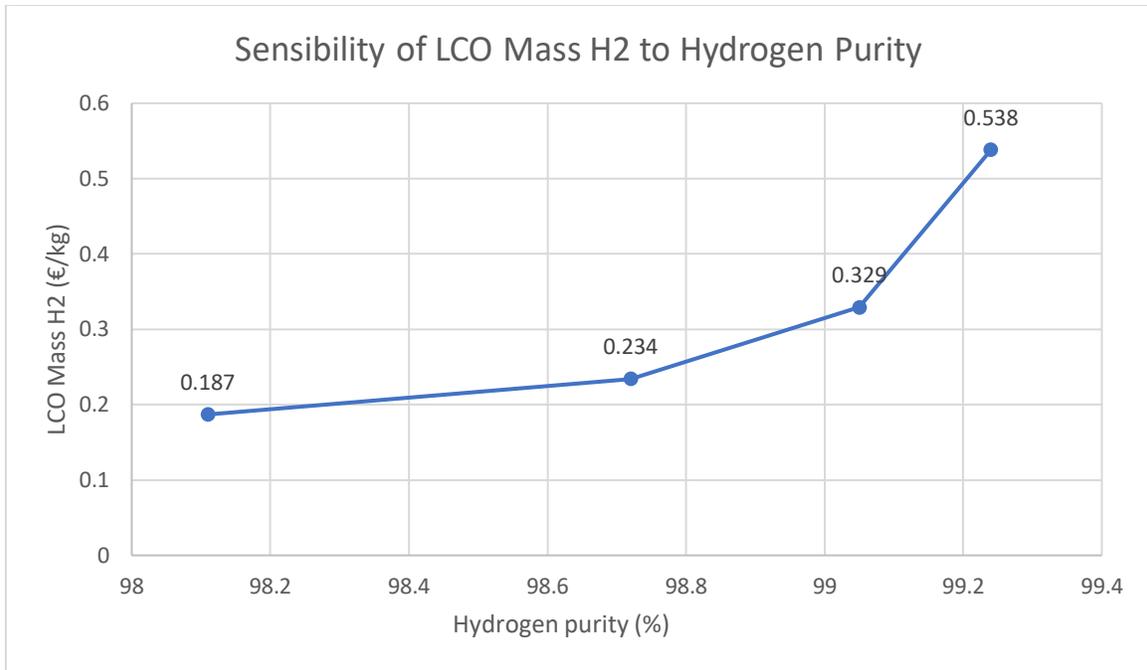


Chart 5 Sensibility of LCO Mass H2 to Hydrogen Recovery

The conclusion on this sensitivity technoeconomic analysis is that from the calculations above we observe that the compressor investment is massive compared to the HX's investment; besides, the costs of electricity to operate the compressor increase proportionally with the size of it. This backs up the premise that we strive for a model where a major fraction of the residual gases (methane) resulting from the separator vessels at every stage must be kept at higher pressures, so the compression work to be done is less and therefore the investment and the operation costs are less.

Chapter 8 Further Sensitivity Analysis of changing the pressure drop at the PRI Case

8.1 50 – 24 bars Case

For the sake of having a sense on the variations of the results if used another representative pressure drop used in Italian grid, we run the model in which the gas is received at 50 bar and the residual gas is delivered at 24 bar.

So, to start the same three cycle selected model setup described before is considered (feed gas composed of 80% mol methane and 20% hydrogen), however now an inlet pressure is 50 bars (transmission grid) and the delivery pressure (distribution grid) requested is 24 bars which again corresponds to an actual representative inlet and outlet pressures used at the PRI points by the Gas Grid operator in Italy. The mass flow corresponds to 1 MCMD.

8.1.1 Model Use Case Results

The internal pressures chosen at every valve are updated to this Case, however most of them follows the same used in the Base Case but updated by this case and are shown in the table below:

Table 15 New Model Internal chosen pressures drops.

Valve		GASPRV	HPPRV0	HPRV1	MPRV	LPRV	HPRV
Name							
Defined inlet pressure [bar]	50	50	30	30	30	30	30
Defined outlet pressure [bar]	30	24	24	12	3	20	
Calculated pressure drop [bar]	20	26	6	18	27	10	

As the differences, it can be noted that in this case the outlet pressure at the LP cycle is 3 bar which is higher than the 1.5 bar of the previous case and also that the delivery pressure required for the Transmission Grid is 24 atm which is higher than the 5 atm of the previous Case (suggesting that higher amount of compression work will be needed).

In fact, from the results of the model, it is noted the corresponding set of results:

Table 16 Product (Hydrogen) Results main metrics New model

Hydrogen recovery	90.2%
Hydrogen molar flow (kmol/hr)	308.85
Hydrogen mass flow (kg/hr)	617.7
Purity Hydrogen (%)	98,11
Comp Work (kW)	854.81
Total HX'area (m2)	412

Although the Compression Work is higher as expected, the HX's area is higher to achieve the same level of cooling (since less cooling is achieved due to the expansions being down to higher pressures). Therefore, higher areas to exchange heat were required at the HX.

In terms of its LCOH, the cost of the compressors is the most determining factor in the LCOH calculation (HX's are almost negligible compared to the compressor cost and the rest costs are proportional to the compressor investment) and therefore the LCOH is higher compared to the previous case reaching 0.417 (€/kg).

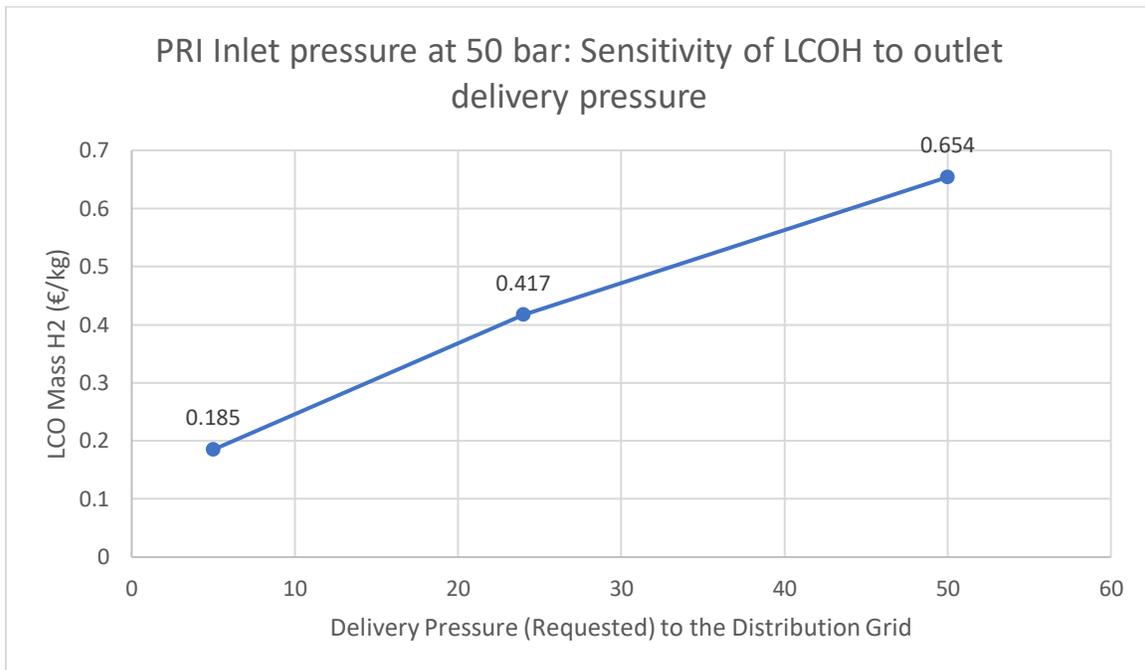


Chart 6 PRI Inlet pressure at 50 bar: Sensitivity of LCOH to outlet delivery pressure

It is noticeable that in this case the necessary cooling to produce hydrogen in the cryogenic configuration was achieved even when increasing the outlet pressure at the 3rd cycle (by means of higher heat exchanger area), however still it is more costly. This is because of the requirement to deliver the feed gas at higher pressures. In fact, the LCOH linked to the requirement for delivery at 50 bar, that is the gas is received at 50 bar and the residual gas delivered at 50 bar, confirms this as it can be seen in Chart 6. Therefore, the primary cause of the increment in the LCOH is related to the higher compression work associated to the need of recompression to meet a certain delivery pressure.

This also suggests that in the case of a PRI where the inlet feed pressure is at 24 bars and the delivery pressure is at 5 bars, the necessary cooling is achieved as in the first Case (by means of adjusting heat exchangers) and the LCOH will remain similar to the first case (same necessary compression work to bring the residual gasses to 5 bars).

8.2 Sensitivity analysis of LCOH

For the following cases the sensitivity analysis is carried out over the original Model Use Case used in this report. That is the Case 50-5 and following the values shown in Table 1 and Table 2.

8.2.1 Sensitivity of LCOH to electricity price

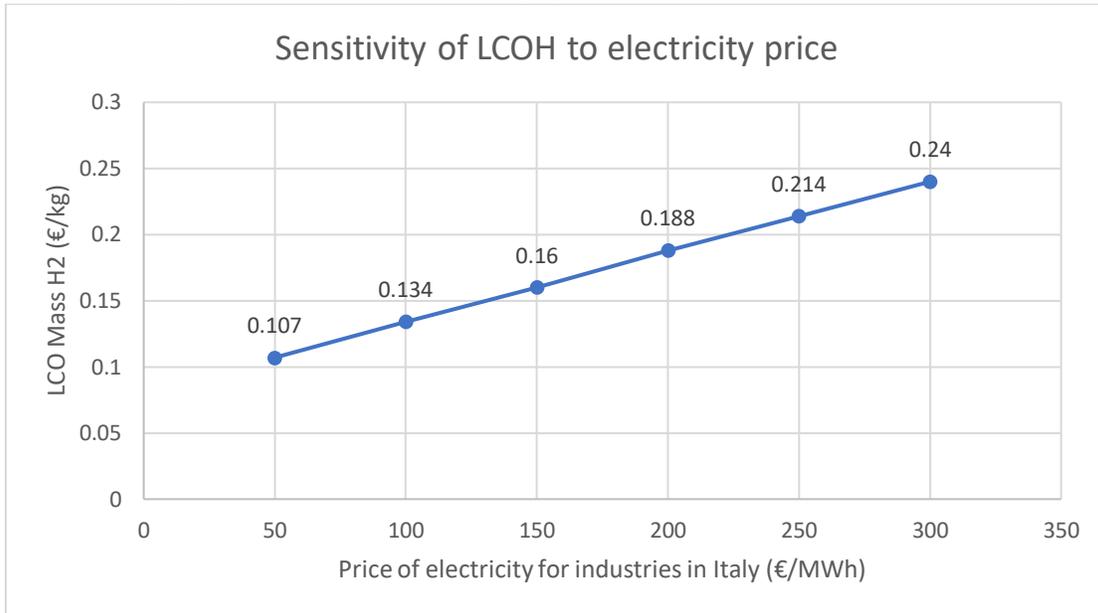


Chart 7 Sensitivity of LCOH to electricity price

It is observed that the influence on the LCOH to the variation of the price of electricity is minimal. In fact, it takes 100 euros increment in the price of electricity per MWh to only increase by 5 cents the price of a kilogram of hydrogen deblended.

8.2.2 Sensitivity of LCOH to the number of hours of functioning of the deblender with an Uncertainty Factor of 80%

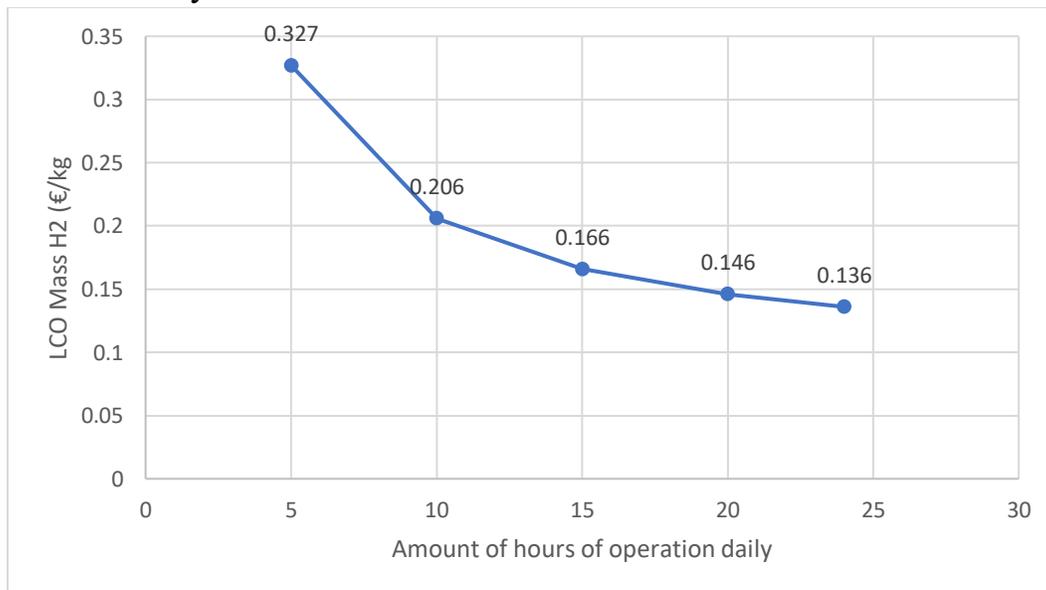


Chart 8 Sensitivity of LCO to the number of hours of functioning of the deblender (with an 80% UF)

Regarding the amount of time that the facility is operating it is observed that as the operating hours in a day increase the LCOH decreases. The way it does it is asymptotically suggesting that from the chosen 15 hours a day it is not much convenient to further increase its use as the decrease in the LCOH starts becoming increasingly less. Likewise, under operating the facility translates in increasingly higher LCOH.

Chapter 9 Conclusion

The objective of attaining a rigorous evaluation on the scope of deploying deblending equipment in the context of blended hydrogen-methane distribution has been accomplished. In fact, the results are promising, and it is encouraging to the political measure of choosing this blended distribution scenario to develop and foster the hydrogen market and use.

From the technical point of view the simulation showed the inverse relation between the HX's area required, and the compressor power needed. That is, the further the feedgas is expanded down in pressures; and consequently, the power to recompress afterwards increase, the lower HX's area will be needed. This initially gives the designer of the facility full freedom to choose among these variables the one to let go high and the other low as in both cases the cryogenic targets are met. However, the technoeconomic analysis strongly tells the designer to prioritize in leaving the recompression power need the lowest as possible and let the HX's are go high since the cost of the first is many folds more costly than the later.

In the case where the cooling target is let free to vary as a function to the level of expansion, the higher purity related to the latter shows to be unjustified considering the massive increase in the LCOH associated.

In the overall analysis the fact that highlight from the technoeconomic evaluation is the low LCOH to deblend hydrogen obtained. Even though the LCOH behaves increasing elastically as the delivery pressure required is set higher, this still reaches manageable costs as it is shown in the case of having to deliver at the same pressure that it is received (50 – 50 bars). In that case the LCOH reaches up to 0.654 €/kg which is a reasonable price considering that it represents just 20% of the medium-term projected cost of the kilogram of green hydrogen produced by electrolysis (2.5 €/kg).

9.1 General Context for Deblending technology deployment

Deblending facilities deployment possibility comprises 2 main processes as outlined below.

9.1.1 Modelling of a deblending machine (processes developed in the present report).

The first of these processes consists in modeling in Chemical engineering Software the deblending equipment (machine) able to separate hydrogen or methane from a methane-hydrogen mixture using cryogenic separation method, which is one of the two main separators methods employed along with membrane separator.

Using the Software, the aim is to obtain the compelling, relevant values and metrics that are useful to know the performance for the different nodes where a deblending equipment is deployed.

Afterwards, an implementation phase (trials) at the laboratory is needed to estimate the accuracy of these modelation and to be the previous step to set it in practice.

This process continues assigning the closest standardized (proofed) use case to the specific case since a significant number of network locations operate at the representative conditions selected. Consequently, all the relevant specifics metrics of the system will be output (National, 2021). Once we have this, the selection of the separator technology needs to take place.

The whole analysis to get to these inputs metrics that will output the equipment technology to be installed in the project and its cost will be determined by the assessment that identify a specific project requirement, drivers, utility available on site, land availability, environmental and planning application, etc;

consequently, integrating all these technical-wise factors to the process of selection of the equipment and technology.

Because grids vary significantly from one area to another, and this is specially the case when a vast area is being considered for the application of this solution such as a continent or an entire country, each and every potential point for deblending equipment deployment needs to be evaluated as it can affect the variables that were outlined above impacting the decision.

9.1.2 Social, political, legal, and economic considerations related to its site implementation selection.

The second of these processes delineates the selection of deblending facilities sites which is triggered mainly by commercial considerations of having all the clients' specific uses requirements; that is, each client would request a certain product need in the context of deblending technology operating scenario. This evaluation by the operator can be seen as a function that need to be optimized which is maximizing the benefits to all the actors and has for variables restrictions the legal agreed blending ratio into the grid (for example 20% mol H₂), the chance to suggest adaptation of equipment to the clients (for example adapt it to tolerate a certain mol% of Hydrogen in the blend for hydrogen sensitive users), aggregating the most possible clients together that can benefit from the specific place of the facility, among others. For this, an extensive data acquirement will have to be obtained from all the actors (clients, operators, auditors) with the aim to define a decision-making procedure that will draw the main guidelines for the operator to proceed which will be both quantitative and qualitative in nature taking all these variables into account.

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