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Techno-economic assessment of manure to hydrogen plants

A real case study based on a cattle farm in an isolated location

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ABSTRACT

The energy transition is of utmost importance to tackle climate change. Different strategies have been put in place with the objective of reducing greenhouse gas emissions, increasing the share of renewable energy (RE) or energy efficiency. In this scenario, hydrogen has been recognized as a promising energy vector due. However, while hydrogen can be produced in different ways, only some of them are zero emission. This includes green electrolysis as one of the most consolidated alternatives due to the potentially low levelized cost of hydrogen (LCOH) obtained, but other sustainable methods are also promising which includes the reforming of biogas from manure or biomass.

There are two main configurations of manure to hydrogen plants. In the first case, there is a reformer reactor, low and high temperature shift reactor and a pressure swing adsorption (PSA) unit for hydrogen separation. In the second configuration biomethane from biogas upgrading can be the source of hydrogen production, but in case of different necessities biomethane is available as a by-product too.

In order to study the potential of manure-to-hydrogen plants with the two configurations described, a case study based on a real location has been assessed in this thesis based on the FCH JU REMOTE project in a demo-site located in La Aldea de San Nicolás, Gran Canaria, Spain, where there is a plan to consolidate a hydrogen valley in the medium to long term.

Thus, a techno-economic analysis has been performed to assess the two configurations described above. Results show lower CAPEX and levelized cost of hydrogen (LCOH) for this plant and the input manure stream in the location.

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1 INTRODUCTION AND OBJECTIVES

1.1 Introduction and rationale

Year by year, global energy consumption grows due to continuously increasing services request by human population, in the last century energy request rose especially for developed countries and consequences are visible day by day.

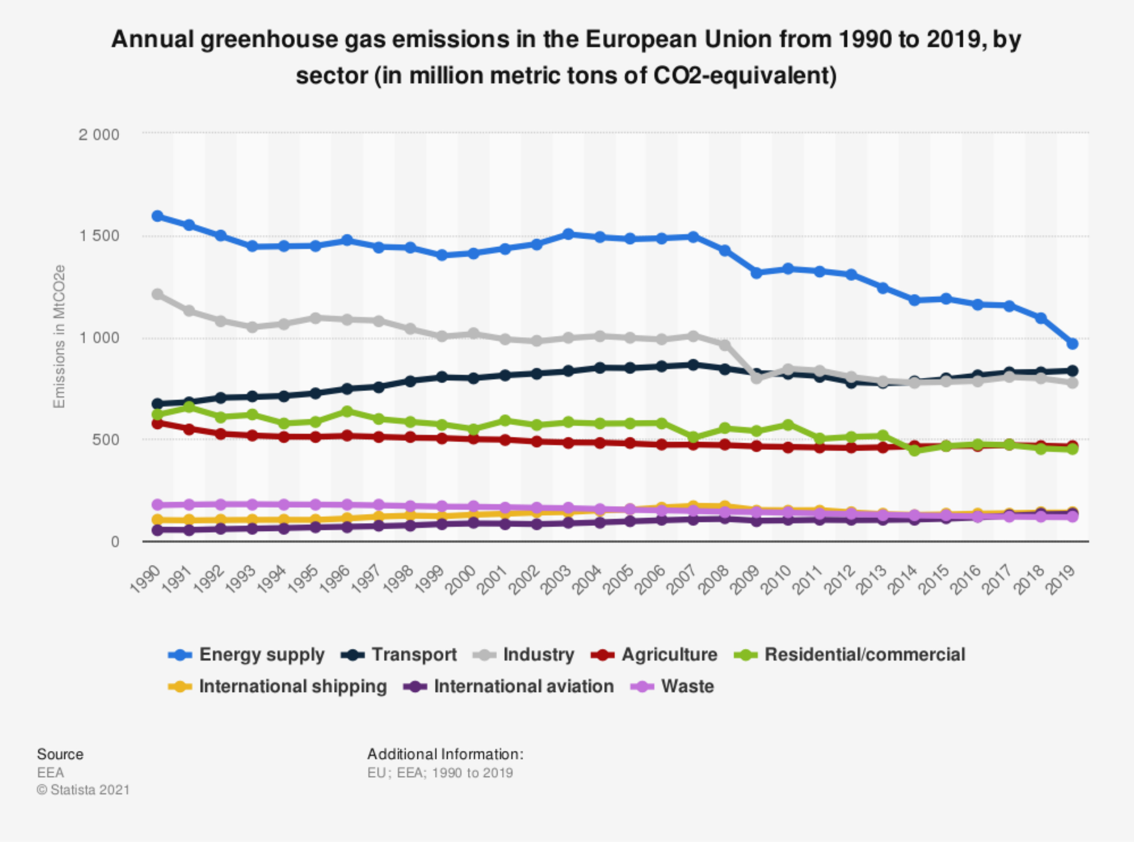

Figure 1 - Annual greenhuose gas emission in the European Union (1990-2019) [1]

Most of the energy is still provided to the end users by means of fossil fuels and consumable resources that must be extract from Earth: such activities provoked destruction of natural habitats, eco-systems pollution, animals and plants extinction and social injustices. The connection between industrialization and global warming is, nowadays, very well known: the use of fossil fuels has led to an increase of the CO_2 percentage in the atmosphere. Solar energy from the sun is naturally released in space by the Earth, but thanks to the atmosphere and greenhouse gases, some of this energy remains trapped here: with GHGs presence the Earth surface is maintained at the right temperature for human, animal and vegetal species to live on. With industrialization, greenhouse gas emissions increased bringing CO_2 presence in atmosphere from 280 ± 10 ppm before 18th century, to 367 ppm in 1999 [2] and 419 ppm in 2021 at Mauna Loa Observatory [3] [4][5].

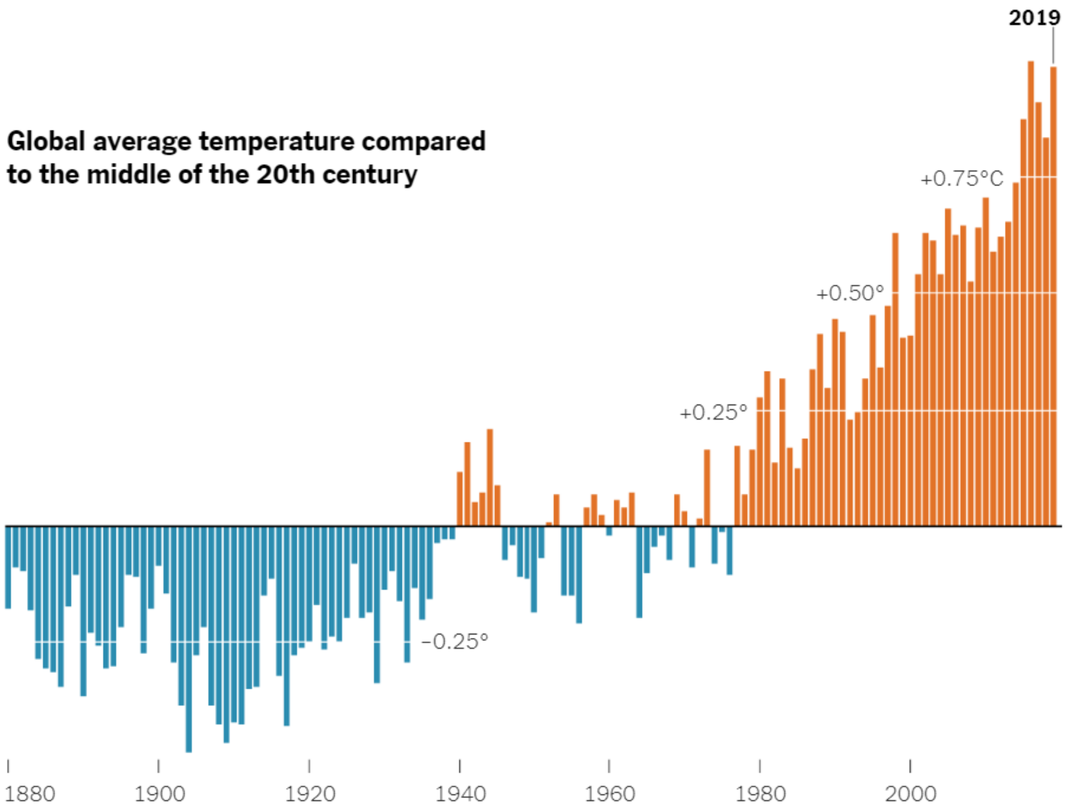

Figure 2 - Global average temperature compared the middle of the 20th century from New York Times

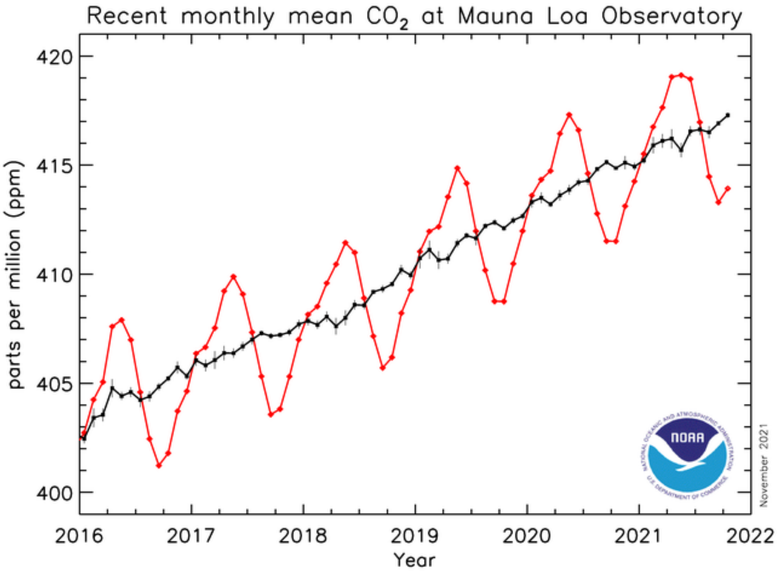

Figure 3 - CO_2 concentration at Mauna Loa Observatory of the last years

Evidence of global warning are now clear: global average temperature increase is leading to irreversible ice melting Arctic, Antarctic and glaciers regions: water flows, then, in seas and oceans increasing their levels. According to "Climate Change: Evidence & Causes" by Royal Society, global warming contributes to water evaporation on land that is increasing extreme weather conditions such as droughts, wildfires and hurricanes [4].

Additionally, sea level rise will affect all the populations living in coastal areas: IPCC reports that, according to different kind of simulations, mean seas level rise can reach around 0.26 mm/year in the period 2014-2065 and 0.60 mm/year in 2100 [6]. Numerous projections had been simulated for different scenarios (from the most realistic to the most pessimistic one) and over the 21st century in case of low emissions an increase of about 0.29-0.59 m and around 1.5-2.5 m for upper bounds [7]. Even though it may not seem a high increment over a century, it's important to consider that islands (such as Maldives) or cities (such as Venice) would be completely flooded and irreparably lost.

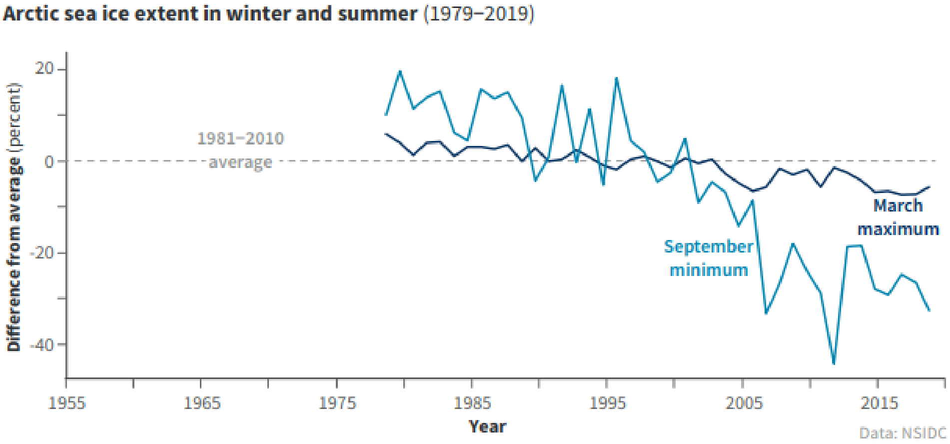

Figure 4 - Arctic sea ice in winter and summer (from 1979 to 2019) [4]

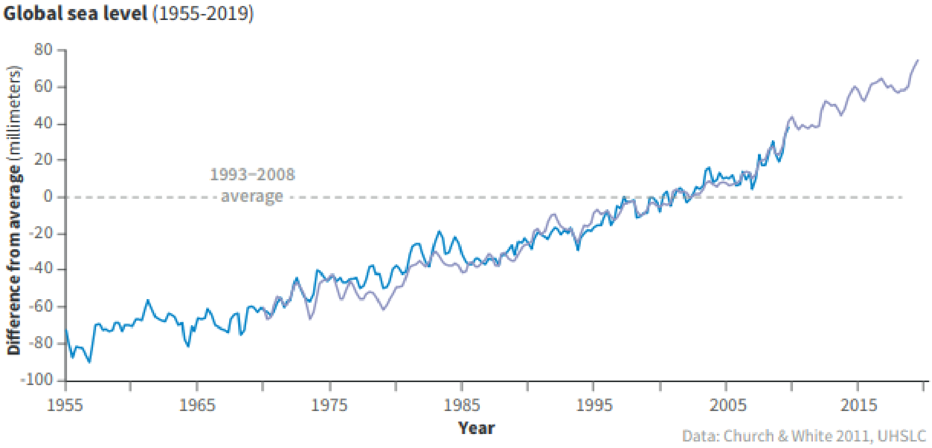

Figure 5 - Global sea level (from 1955 to 2019) [4]

Even though a lot of attention is put in carbon dioxide presence in atmosphere, global warming is also linked to other emissions: other GHGs (such as methane, nitrous oxide and halocarbons) are now emitted during industrial processes, animal breeding in farms, heating, transportation and other human activities [4]. According to the reference previously cited [4], the carbon dioxide concentration in the atmosphere increased of 40% (always compared to preindustrial levels, methane of 150% and nitrous oxide of 20% [4]. Carbon dioxide is a very stable molecule and its lifetime is quite long: every CO_2

molecule has a lifetime between 5 to 200 years, for this reason it is fundamental to reduce emissions [8].

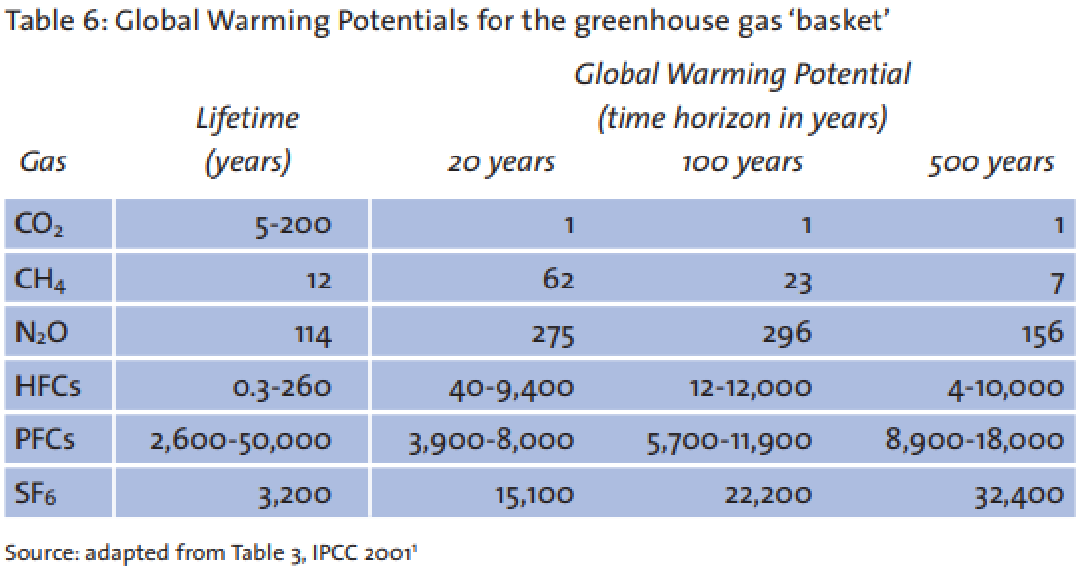

Table 6: Global Warming Potentials for the greenhouse gas 'basket'

Gas	Lifetime (years)	Global Warming Potential (time horizon in years) 20 years	100 years	500 years
CO_2	5-200	1	1	1
CH_4	12	62	23	7
N_2O	114	275	296	156
HFCs	0.3-260	40-9,400	12-12,000	4-10,000
PFCs	2,600-50,000	3,900-8,000	5,700-11,900	8,900-18,000
SF_6	3,200	15,100	22,200	32,400

Source: adapted from Table 3, IPCC 2001[1]

Table 1 - Global Warming Potential and GHGs lifetime [8]

Lately, technical development of renewable energy technologies allows achieving higher efficiencies and increase the energy conversion. European Union and United Nations have defined several goals to support energy transition combined with financial funding for projects that aim at decarbonization and energy efficiency. Below, some of the most important agreements in this matter are summarized:

- **Paris Agreement.** During COP 21, Paris Agreement have been signed by the participant nations: such agreement has the goal to help the decarbonization process in order to hold the global average temperature below 2°C (and possibly below 1.5°C) above the pre-industrial levels [9], [10]. Additionally to this macro-goal, smaller objectives had been defined for nations: reduction of greenhouse gas presence in the atmosphere, increase of the amount of energy produced with renewable energy sources and increment in the consumption optimization in order to reduce energy and material wastes. To do so, financial structure has been developed, through it funds and subsidizes have been given to companies to help them develop new technologies.
- **Green Deal.** European Union had set a very important milestone for countries: 20% emission reduction of greenhouse gases (GHG), 20% of energy produced from renewables and 20% improvement in energy efficiency [11]. Since the positive impact of these goals, new targets have been set for 2030 and 2050:
    - Greenhouse gas emission reduction to at least 55% by 2030 with respect to 1990 levels

  - No emissions by 2050 in order to make Europe the first climate neutral continent [12]
- **COP 26.** During the writing of the present work, one of the most important event about renewable energies and climate change occurred: the Conference of the Parties (COP 26) took place in Glasgow from 31 October to the 12 November of 2021. During COP 26 nations discussed about climate change and energy transition to define goals and new pathways to protect and restore ecosystems, finance the acceleration of the decarbonization process maintaining the medium temperature increase below 1.5°C [13], [14]

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  The conference has produced the Glasgow Climate Pact, in which urgency and concern due to human activities is expressed by the Intergovernmental Panel. The Conference of Parties recognizes the increase of the average global temperature of around 1.1°C since the pre-industrial levels and emphasizes the necessity to increment financial support for climate change. To all Parties is requested more commitment than ever and the integration of climate supporting and energy transitioning strategies.

  In the “Mitigation” section of Glasgow Climate Pact, the Conference of Parties points out the necessity of reducing the global greenhouse gas emission by 45% by 2030 with respect to 2010 levels and obtaining net-zero carbon emissions by mid-century. Methane is also a greenhouse gas and COP 26 recommends to control and reduce its emissions [14].

  A certain attention is also given to the financial aspects: fundings and investments will be unlocked for clean energy technologies and support energy transitioning process. At COP 26, has been also reported that the USD 100 billion per year to for developing countries have not been fully delivered by 2020 and a new deadline for 2025 is set. Thanks to the continuous support, countries under development will be able to decarbonize faster while their development continues.

  Specific emphasis to the role of indigenous people in preserving local communities’ culture, knowledge and biodiversity helping action on climate change and appreciation is expressed for youth non-governmental organizations that fight for the cause [14].

> A last-minute variation proposed by India was quite controversial and opened an interesting debate: Indian delegation forced the Conference of Parties to change in the draft document the term “phase out” with the less compelling “phase down”. India affirms not to be ready for coal neglection for the next years because the country still needs to develop [15], [16].

In this context, the potential of hydrogen technologies resides in accommodating the surplus renewable energy produced in a context with a high penetration of these energy sources. In particular, hydrogen is an energy vector suitable for storage of large amounts of energy which can be produced, stored, transported and distributed in many ways, with a wide range of end uses [17], [18].

However, the problem of producing green hydrogen resides in the high CAPEX of the technologies as well as their operational costs. In this context, several funding instruments have been activated in the latest years to foster technology development in the EU. Horizon2020 is a funding program that aim at sponsoring green project and ideas that can help the energy transition: 80 billion euros have been distributed in seven years (from 2014 to 2020) [19]. Among the programs, the FCH JU is the public-private partnership funding R&D and innovation activities around hydrogen. Among the financed projects, REMOTE is one of them and it includes 3 DEMO plants.

1.1.1 FCH JU REMOTE Project

REMOTE (Remote area Energy supply with Multiple Options for integrated hydrogen-based Technologies) is an EU-funded project aimed at the demonstration of the technical and economic feasibility of a fuel cell-based H_2 energy storage solution. Some DEMO plants have already been installed in selected areas: one is operative in Froan Island (Norway), one in Agkistro (Greece) and the last one is under construction in Gran Canaria (Canary Islands, Spain).

The main goal of the project is to provide maximize the renewable energy resources in remote areas via power-to-power solutions based on hydrogen. Renewable energy comes from fluctuant and unpredictable resources and, consequently energy loads cannot be always covered. Storage systems are installed to reduce the effect of irregularity in the production for the end-user. Today, in remote areas, two main technologies are used to store energy: electrochemical batteries and hydrogen storage.

This master thesis is based on Canary Island case study: a farm owned by Grupo Capisa is located on Gran Canaria, on the western part of the island, near La Aldea da San Nicolas. The whole municipality is considered as Protected Rural Soil and it’s surrounded by three national parks (Natural Reserve of Güi-Güi, Natural Park of Tamadaba and

Natural Park of El Nubio). The electrical grid of the island cannot be built through these parks and the farm is currently using diesel engines to power the entire complex. The farm is a milking facility and it is increasing its cattle population from 750 to 1,000 and it needs 40 kW that covers part of the load; an additional 100-120 kW peaks needs to be covered to cover the new electrical loads that are now under construction.

A side section has been considered for future development: a conversion line can transform manure into biogas and, lately, hydrogen. This project brings a renewable and constantly produced source (manure) from waste to useful energy. In this scenario, the additional hydrogen flow can power a hydrogen refueling station in town to charge fuel cell cars, buses and trucks passing through La Aldea: such process helps in decarbonization of the transport sector on the island. The final goal is to deploy a hydrogen valley in a remote area where upgrading of the electricity infrastructure is not evident.

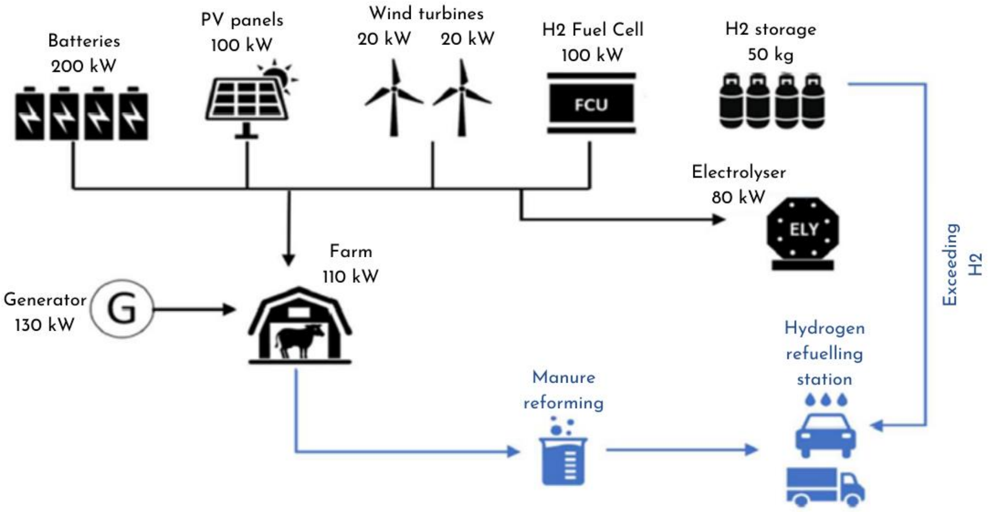


Figure 6 - FCH JU REMOTE Project coupled with La Aldea del Hidrógeno

1.2 Objectives

The main objective of the thesis consists in the production of a techno-economic analysis of a manure-to-hydrogen plant planned be the continuation of the REMOTE European project in the demonstrator in Gran Canaria, Canary Islands, Spain.

The analysis starts from the evaluation of the feedstock potential from 1,000 livestock of the farm, where manure is daily produced and it can be converted by means of an anaerobic digester into biogas and digestate. A fundamental part of the thesis consists in the assessment of the biogas produced and its successive purification and storage. Aspen Plus is used for the design of the hydrogen production phase: steam methane reaction (SMR) and water gas shift (WGS) occur and break the bonds of methane and water to free hydrogen. Sensitivity analysis on steam-to-carbon (S/C) ratio are performed in Aspen Plus to evaluate the optimal amount of steam that should be sent to the SMR reactor to have high conversion rates and avoid carbon deposition. The technical design is followed by the economic assessment of the plant, the cash flow analysis and the calculations of levelized cost of energy and hydrogen (LCOE and LCOH). Economic feasibility is later on supported by the estimation of town fleet that can be hydrogen fueled thanks to the project.

To reach the main objective of the thesis, the following particular objectives are pursued:

- Evaluation of cattle feedstock potential
- Biogas production quantitative evaluation
- Multiple biogas purification analysis
- Biomethane alternative conversion path
- Technical plant design (from manure to hydrogen)
- Design of the hydrogen production phase in Aspen Plus
- Sensitivity analysis on S/C ratio in Aspen Plus
- Economic and cash flow analysis
- Levelized Cost of Hydrogen and Energy (LCOH and LCOE)
- Hydrogen consumption and final use assessment

1.3 Scope

Below, the scope of the thesis is presented in blue color over the black blocks which are relative to the current activities in REMOTE. As it can be observed, the target is to add a hydrogen stream to the production available from renewable energy via de use of cattle manure, maximizing the potential of the farm facilities in a concept going towards the idea of circular economy.

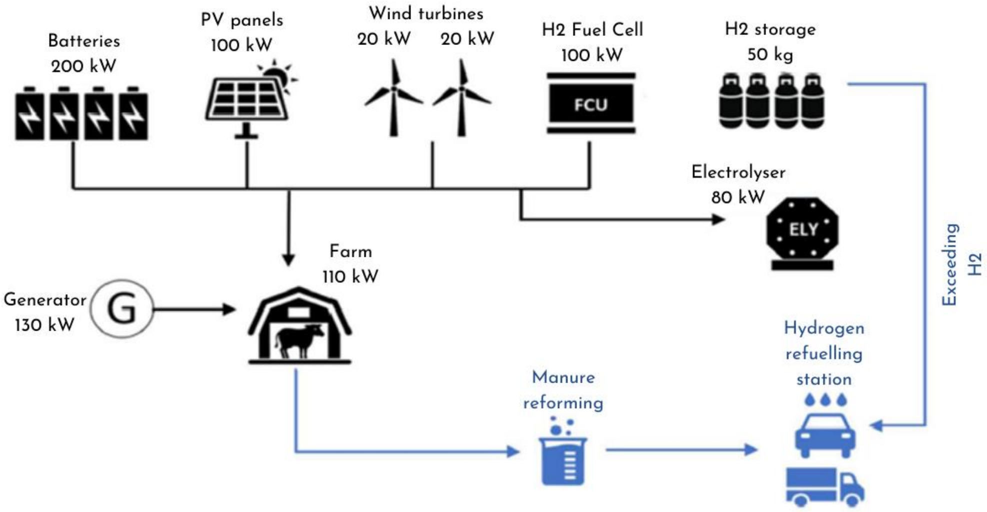


Figure 7 - FCH JU REMOTE Project coupled with La Aldea del Hidrógeno

1.4 Contributions

The present work goal is to assess the techno-economic feasibility the hydrogen production facility starting from manure: in the hydrogen generation panorama is explored as an alternative pathway since renewable and biological source is used. The novelty in the work consists in the coupling of two well-experienced technologies such as anaerobic digestion and steam methane reforming. The project will contribute to powering the hydrogen economy at La Aldea de San Nicolas push and accelerate the hydrogen economy development in the area allowing the town in achieving the energy independence.

1.5 Contents and structure

The present work is developed around the following sections:

- Part 2 explains the plant design, with a brief description of the equipment, technologies and physical fundamentals that drive the reactions and working principles of the designed system. The technical and economical comparison of the different pieces of equipment is carried out.
- Part 3 describes the case study of La Aldea and the approach adopted for the techno-economic analysis of the following chapters
- Part 4 contains the technical evaluations for the feasibility evaluation of the plant, the sensitivity analysis and the power consumption study. Aspen Plus® software has been used to design and simulate the hydrogen production section: sensitivity analysis on S/C ratio at the inlet of the steam methane reformer has been performed to evaluate the effects of the steam quantity in the reactor. The software provides heating and cooling power consumption for every single equipment present in the flowsheet, combined with literature data on energy consumption (necessary for the other components of the plant), the whole energy consumption is obtained. Part 5 describes the economic analysis for the evaluation of CAPEX and OPEX. Financial feasibility is determined by LCOH, LCOE, IRR and payback time. Depending on the values obtained, in-depth financial evaluations are reported to define the economic feasibility or unfeasibility of the project. Some observations and considerations are also reported on a hypothetical FCEVs fleet that can be powered by the plant: cars, buses, forklifts and ships can be considered

2 STATE OF THE ART

This section presents the state of art of technologies and processes applied in the work done, as well as the assumptions underpinning it. Below, the topics subject to research in order to develop the case study in the thesis are summarized:

- Configurations of manure-to-hydrogen plants
- Biogas, biomethane and hydrogen production
- Anaerobic digestion
- Biogas purification
- Biogas upgrading
- Biogas and biomethane storage
- Steam methane reforming (SMR)
- Dry reforming
- Water gas shift (WGS)
- Hydrogen purification
- Hydrogen compression and storage
- Hydrogen trailers
- Hydrogen refueling stations (HRS)

The following subsections provide the state of art and the techno-economic overview for each of these domains which are used in the case study further developed.

2.1 Configurations of manure-to-hydrogen plants

The first main configuration observed in the state of the art consists of hydrogen production from biogas previously obtained by anaerobic digestion [20]–[22]. Such concept can be developed in locations with high manure availability, manure is converted in biogas with an anaerobic digester. Once biogas is purified, it can be compressed and stored in a buffer tank and lately used in a hydrogen production plant, with steam methane reforming, dry reforming and water gas shift, hydrogen is obtained. Lastly, hydrogen is compressed, stored in a buffer tank and used in different applications. In the figure below a scheme of the plant is reported for the case of mobility with hydrogen vehicles:

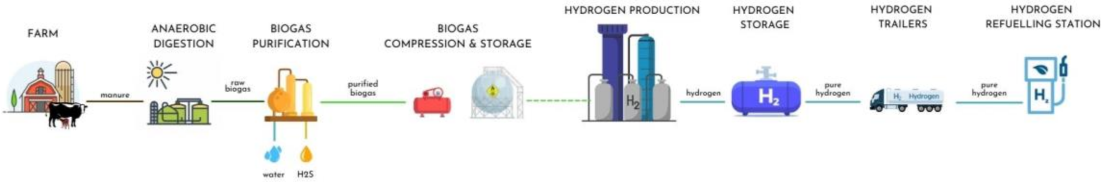


Figure 8 - Plant layout of the first configuration

A second configuration that introduces biomethane is also prevalent in manure-to-hydrogen facilities [22]–[24]. Such case has the same concept of the first configuration, but a biogas upgrader is added and biomethane is produced. In the hydrogen production section, steam methane reaction and water gas shift is applied only, and it send lately to the hydrogen refueling station.

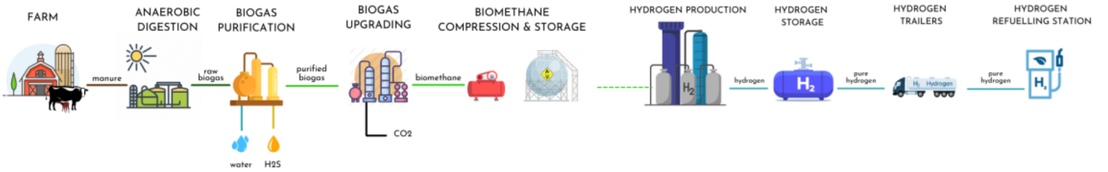


Figure 9 - Plant layout of the second configuration

Such technology is also very promising from the carbon capture utilization and storage (CCUS) point of view: the system is able to efficiently separate carbon dioxide from hydrogen and other gases. In the present work such technology has not been introduced in the plant, but certainly it represents a possibility for further analysis and observations.

2.2 Anaerobic digestion

Anaerobic digestion is a biological process that occurs in oxygen absence: it consists in the transformation of the organic substance contained in manure into biogas. Biogas is mainly a mixture of methane (CH_4), carbon dioxide (CO_2) and other species such as hydrogen sulfide (H_2S), carbon monoxide (CO) and water (H_2O).

The degradation process is composed of four main phases: hydrolysis, acidogenesis, acetogenesis and methanogenesis [25].

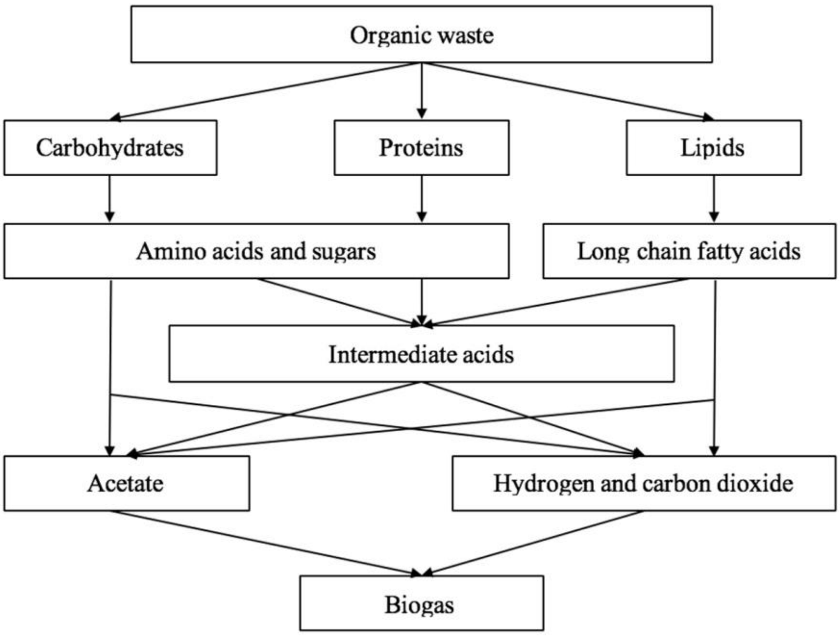


Figure 10 - Simplified scheme of pahway in anaerobic digestion [25]

Organic biomasses used in anaerobic digestors usually are characterized by complex polymers, through hydrolysis these molecules can be transformed into smaller ones

Hydrolytic bacteria use enzymes to turn carbohydrates, proteins and lipids into sugar, amino acids and long chain fatty acids, respectively. For particularly complex substrate, like lignin and hemicellulose, may need pretreatments or additional enzymes to enhance the degradation process.

Usually, optimal temperature for hydrolysis is in the range 30-50°C and optimal pH is between 5 and 7.

Hydrolysis is the rate-determining step and it influences the kinetic of the reaction, it's the bottleneck of the due to its slow rate of conversion.

Some strategies to improve hydrolysis can be adopted: co-digestion with C-rich substrates, thermophilic hydrolysis and stage separation.

Acidogenesis follows: acidogenic microorganisms absorb hydrolysis products through their cell membranes and convert it into intermediate volatile fatty acids (VFA). This process is much faster than the others because acidogenic bacteria regenerate in less than 36 hours. Ammonia is produced by deamination during the process and, if not controlled, it can inhibit anaerobic digestion.

The third reaction consists in the transformation of VFAs into acetates and hydrogen and it's called acetogenesis.

Methanogenesis is the last step, methanogenic microorganisms convert acetates, hydrogen and carbon dioxide into biogas. Higher pH values are required with respect to previous stages in order to have optimal reaction conditions [25].

Generally anaerobic digestion retention time should be between 10-40 days, depending on the source: the real hydraulic retention time (HRT) can be obtained with the type of substrate and the chemical composition of the inlet flow (in our case cattle manure) [26].

Two main temperature operation zones can be defined: mesophilic digestion occurs around 35°C, while thermophilic one around 55°C. Due to lower temperature, mesophilic digestion is a slower process than thermophilic digestion, also biogas yield is lower. Psychrophilic digestion needs temperature lower than 20°C, but they have less applications [25].

Anaerobic digestor specific investment is 200 €/m^3 [27] according to M. Salerno et al.: this cost comprehends the anaerobic digestor (no specific model or technology is specified) and all the supporting components and auxiliary systems necessary.

Also, operational costs are reported in the paper, they are divided into full service (maintenance included), insurance policies, biology monitoring and digestate scattering [27]:

OPEX **CALCULATIONS**		
Insurance policy	3,000	€/y
Biology monitoring	15,000	€/y
Digestate scattering	5,000	€/y
Labor cost	51,250	€/y
OPEX	74,250	€/y

Table 2 - Operational costs of an anaerobic digester

To maintain the perfect thermodynamic conditions for anaerobic digestion to happen, thermal and electrical energy are consumed. Oehmichen et al. report range values of 0.054-0.16 MJ/MJ_{biogas} for process heat and 0.0038-0.0059 kWh/MJ_{biogas} for electricity. For sake of simplicity medium values have been used to make calculations:

- Specific energy for process heat = 0.107 MJ/MJ_{biogas}
- Specific energy for electricity = 0.0045 MJ/MJ_{biogas}

2.3 Biogas purification

Once biogas is produced by means of anaerobic digestion, it must be purified by contaminants such as hydrogen sulfide, ammonia and water. Regardless biogas use, this step is fundamental for reduction of substitutions and maintenance costs: hydrogen sulfide can easily react with water corroding tubes and every component the flow encounters.

2.3.1 Removal of water

First step consists in dewatering, biogas drying is necessary and some methods can be evaluated:

- Condensation: biogas temperature is lowered so that all water vapor is removed
- Physical adsorption: silica gel or other elements with desiccant properties
- Pressure increases desorption: relative humidity of the gas is reduced without removing water
- Chemical drying with glycol adsorption

Glycol is the drying agent that is used in chemical adsorption: it reduces absorbs water and it is regenerated later on at 200°C. With respect to physical adsorption, it has higher efficiency in water separation: dew point can't be reduced over 0.5°C (otherwise water will freeze), but by using chemical separation with glycol, the dew point is reduced to -15°C [28].

2.3.2 Removal of H_2S

Second step is desulphurization: H_2S is a toxic molecule for humans and has corrosive properties, maximum attention should be put on this section. Some techniques can be used for desulphurization:

- Chemical adsorption: adsorption through activated carbons
- Physical absorption: water scrubbing
- Air dosing in anaerobic digester
- Membrane separation

Two adsorption techniques are defined: physical adsorption relies on physical attraction forces (Van der Waals and intermolecular bonds), while chemical adsorption relies on chemical bonds, much stronger than the physical ones [29].

Adsorption through activated carbons rely on high internal porosity, pollutants (H_2S in this case) is contained in the desiccant material by means of intermolecular forces. This technique is mostly used for low H_2S concentration request, activated carbon is the catalytic surface for hydrogen sulfide oxidation: the reaction produces elemental sulfur and water and it is favored at 7-8 bar (and temperatures of 50-70°C, reached by compression heat only) [29]. Additionally, to have continuous operation, two connected water columns can be used: one is used for adsorption and the latter for desorption (material regeneration) [28].

Physical absorption can be performed with water or organic solvents, the main difference relies on the solubility level of hydrogen sulfide in the solvent: in water it's lower while than other solvents, but their cost is much higher.

In-situ H_2S reduction by means of air/oxygen dosing in the digester is another option, aerobic oxidation occurs on the surface of the digestate and reduces the H_2S presence of 80-99%, reaching less than 100 ppm. This desulphurization technique may be challenging: supplying oxygen to the anaerobic digestor may inhibit digestion reactions.

Membrane separation allows to divide pollutants from biogas stream by imposing a partial pressure gradient across the semipermeable membrane: only selected molecules are able to pass through it. High- and low-pressure membrane removal systems are present and can work with multistage separation.

Ammonia should be removed from the raw biogas stream, in case of future combustion it can react with oxygen producing NO_x and contribute to greenhouse gases emission [28].

In "Biogas upgrading: evaluations of methods for H_2S removal" provided by the Danish Technical Institute, some processes have been selected for La Aldea plant. First one is the biological purification by BiogasClean: the company guarantees hydrogen sulfide presence after treatment of less than 10 ppm. Filters cleaning will be necessary and the Quick Sludge Remover (QSR®) provides an easy and fast way to maintain the purification system [30]. Its investment cost accounts for 108,000 € for a plant processing 200 m^3/h.

Siloxa system is provided by Siloxa Engineering AG and allows to remove both H_2S and siloxanes by means of activated carbon desulphurization. The 500 Nm^3/h capacity plant has an investment cost of 70,300 € and operational costs that accounts for 35,500 €/year [30].

The last technology taken into account is the caustic scrubber Sulfurex® by DMT: it has been designed to remove H_2S from 20,000 ppm to 200 ppm. The investment cost for the 500 Nm^3/h purification plant is reported to be 127,000 €, while operational costs are 64,959 € [30].

Type	Original cost
Biological by BiogasClean	108,100 €
Activated carbons by Siloxa	70,300 €
Caustic scrubber Sulfurex	127,000 €

Table 3 - Biogas purification investment cost

2.4 Biogas upgrading

Biogas upgrading consists in the production of biomethane from a biogas stream flow: carbon dioxide must be separated by methane molecule in order to increase purity and heating values. Biomethane flow it's much more valuable than biogas one both from the economic and chemical point of view: it can be used, later on, in more applications such as transportation, grid injection and fuel cells.

Also for this step, a lot of techniques are present, but based on "From biogas to biomethane: a process simulation-based techno-economic comparison of different upgrading technologies in the Italian context" article, three main technologies have been analyzed: Pressure Water Scrubbing (PWS), Hot Potassium Carbonate (HPC) and Pressure Swing Adsorption (PSA). Simulations of this paper are already performed considering an inlet flow purified by H_2S and siloxanes already [31].

2.4.1 Pressure Water Scrubbing

This upgrading technique relies on the different solubility of CO_2 and CH_4: a counter-current water flow is used to absorb carbon dioxide. There are two absorption columns, in the first one the main adsorption occurs, while in the second one water regeneration is performed through air stripping, a pump is present for water recirculation. The main adsorption column is operated at 25°C and 10 bar: to reach these conditions a multistage intercooled compressor is adopted [32]. Methane solubility in water in these thermodynamic conditions is 26 times lower than carbon dioxide [28]. Every compressor has a compression ratio of 3.14 and isentropic and mechanical efficiencies of 80%. An intermediate flash is present in the upgrading system to separate liquid and gaseous fractions, it operates at 5 bar and sends the water to the regeneration column, instead gaseous components are recirculated through the second stage of initial multistage compressor.

Biomethane produced is then dried and compressed to 70 bar.

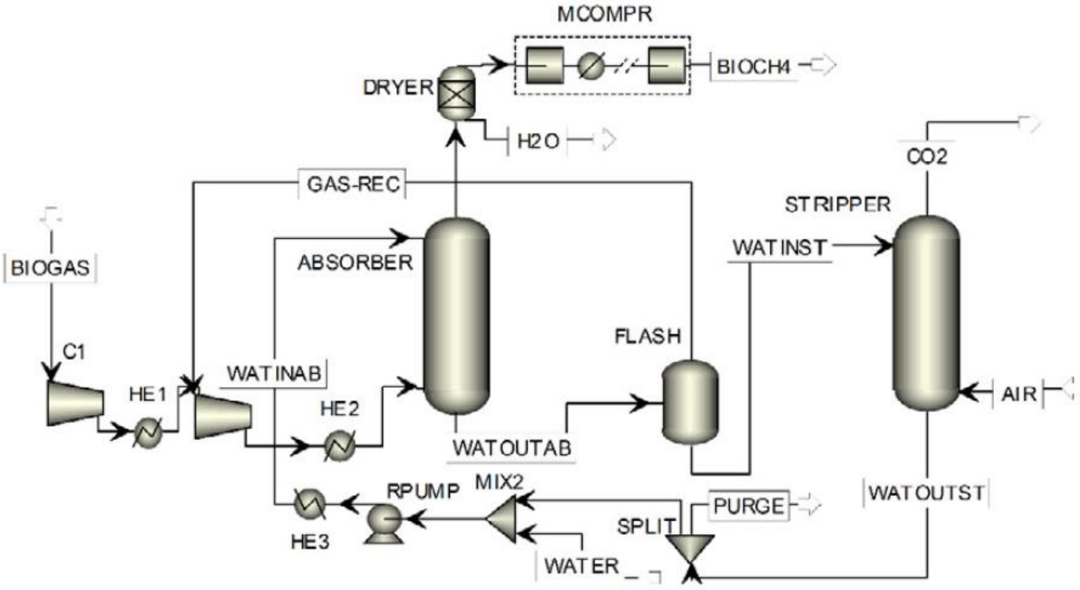


Figure 11 - PWS biogas upgrading layout in Aspen Plus® [31]

For the system, the authors have used Aspen Plus® to model the equipment and simulate the process. A 98.7% purity level is requested as design specification [31].

2.4.2 Hot Potassium Carbonate

Another solution is Hot Potassium Carbonate, it can be used alternatively to amines chemical absorption: it is especially a valuable idea because K2CO3 solution do not bring environmental [33].

Absorption reaction is:

$$K_2CO_3 + CO_2 + H_2O \leftrightarrow 2KHCO_3$$

Equation 1

The systems operate in a very similar way to the Pressure Water Scrubber, but thermodynamical conditions are a bit different: pressure for absorption must be 8 bars and temperature at the top of the column is 70°C. Similarly to PWS configuration, an absorption column is present, followed by a flash separator (for gas that is recirculated and hot potassium carbonate solution that must be regenerated), a second column is present for K_2CO_3 regeneration. Similarly to the previous case, flash operation is set at 5 bar in order to reduce compression ratio and cost; lastly, biomethane is dried and compressed to 70 bar by means of a two-stage compression [31].

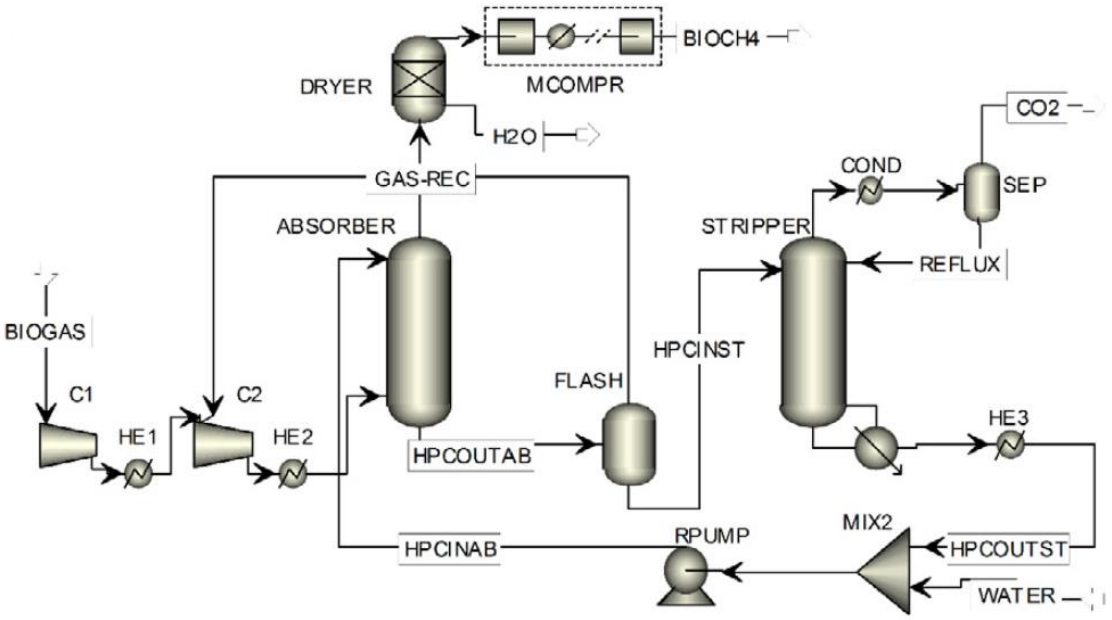


Figure 12 - HPC biogas upgrading layout in Aspen Plus® [31]

The regeneration column works at ambient pressure and 110°C, such temperature is requested to have water in steam conditions enhancing stripping effect and to easily break bonds and regenerate potassium carbonate.

Biomethane purity output of the process is around 98.7%.

2.4.3 Pressure Swing Adsorption

This configuration is quite different from the ones previously described, in figure 13 can be observed six columns (sometimes also four): they are coupled producing three pairs of adsorption/desorption reactions [32].

A cycle of four steps can be observed this technology: pressure build-up, adsorption, depressurization and regeneration. An adsorbent material is used to physically capture CO_2 and the size difference between methane and carbon dioxide particles is exploited for the process. Carbon dioxide molecules are smaller than methane ones, but pressure conditions help the process efficiency.

The adsorption mechanism is performed at 5.8 bar and 33°C with an isothermal process, while during regeneration phase pressure is set at 0.1 bar in vacuum conditions [31].

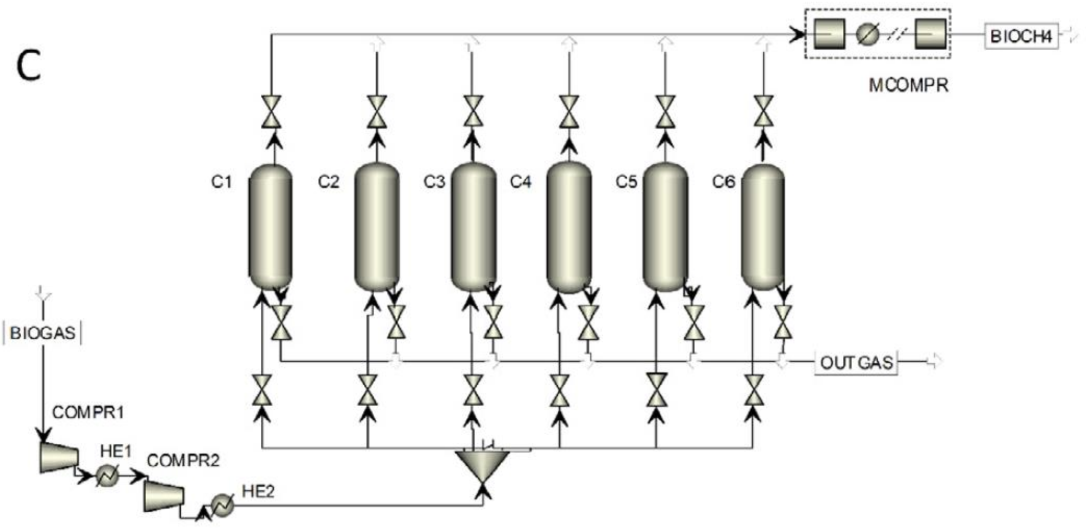


Figure 13 - PSA biogas upgrading layout in Aspen Plus® [31]

A drying system is not necessary because no additional water or humidity is used in the process, so direct multi-stage compression increases pressure until 70 bar.

Typical purity values are reported equal to 98% [32], but a maximum value of 91% is claimed by De Hullu (2008) [31].

In Beccucci's master thesis a biogas upgrading investment cost of 380000 € is reported: such value relates to a 60 m^3/h capacity system, it includes the dispenser and its consumption accounts for 0.4-0.5 kWh/Nm^3 [32]. This system is a high pressure batchwise water scrubbing taken from SGC Report 2013 [34].

With more in-depth analysis, costs for other technologies have been found: in Barbera's paper pressure water scrubbing, hot potassium carbonate and pressure swing adsorption systems are evaluated [31]. In the document 500 Nm^3/h plants are taken into account and their investment costs are evaluated in the following table:

Model	Original cost
Pressure Water Scrubbing	1,323,500 €
Hot Potassium Carbonate (chemical)	1,057,400 €
Pressure Swing Adsorption	1,407,500 €

Table 4 – Biogas upgrading section investment cost

Cost scaling is lately applied to find the right cost for the dimensions of La Aldea plant.

Similarly, operational costs are evaluated from the original values summarized in the following table:

Model	Original cost
Pressure Water Scrubbing	179,581 €
Hot Potassium Carbonate (chemical)	279,094 €
Pressure Swing Adsorption	158,888 €

Table 5 - Biogas upgrading section operational cost

These values are scaled and CAPEX and OPEX are calculated for the economic evaluation. Three technologies are compared: as reported in the graph, the hot potassium carbonate system needs lower investment cost but higher operational costs with respect to the other techniques; figure 14 resumes all the costs:

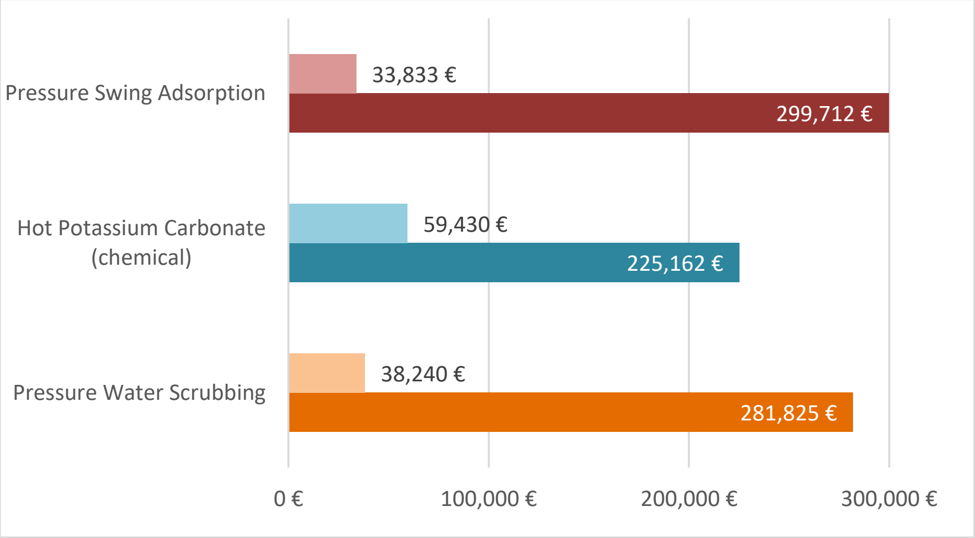


Figure 14 – CAPEX and OPEX comparison for biogas upgrading section

Additionally, the document provides the specific electrical and thermal energy consumption: these values have been obtained in through simulations on Aspen Plus® and are almost equivalent, but it is important to observe that HPC has a high thermal duty because of the regeneration column (regeneration operation temperature = 110°C). Values reported in the paper [31] are resumed in the following table:

Model	Electric Energy (kWh/m^3)	Heating Energy (kWh/m^3)
Pressure Water Scrubbing	0.27	-
Hot Potassium Carbonate (chemical)	0.20	0.67
Pressure Swing Adsorption	0.24	-

Table 6 - Specific electrical and thermal energy consumption associated to different biogas upgrading technologies

2.5 Biogas and biomethane storage

After the biogas purification, and eventually upgrading for second scenario, a storage is added. An accumulation system is necessary and fundamental for this plant and its presence in the design is unavoidable: firstly, it divides biogas production line from the hydrogen one. This characteristic is essential in case of easy and fast maintenance of one side of the plant: with the storage, if the hydrogen section is under maintenance, biogas is still produced and accumulated until the storage is filled. Vice versa, if biogas section is under maintenance, steam reforming and water gas shift reactor are still operative and hydrogen is still supplied.

Storage is able to reduce turn on and off of machineries, reducing costs and energy consumption: biogas supplied to the hydrogen production section can be regulated while it is supplied to the steam methane reforming reactor. Consumption peaks and minimums are absorbed allowing the plant to work on a more stable and controlled rhythm [33].

In general, renewable energies need to be paired with energy storage systems (ESSs) to reduce their intermittency and production oscillations: storages increment plant reliability and productivity [35].

In first scenario, two different type of storages are observed: low-pressure and medium-pressure tank. Low-pressure tanks can be useful to reduce accumulation costs, but need more space; from the other side, medium-pressure tanks can contain more mass in less space and can store more energy, but compression costs must be considered.

Specific costs for low- pressure biogas storage systems are taken from Petrollese and Cocco's paper [36]: it is 40 €/m^3. Medium-pressure storage (at 7 bar) specific cost is higher due to compression: investment is equal to 90.55 $ in Wattanasilp's paper [37].

In second case, biomethane is a more refined product than biogas: biomethane composition mainly consists in CH_4 and a small quantity (around 2% of CO_2). Eventually, biomethane stored can be used for other purposes, depending on the necessities of the farm it could be used to power the steam reforming boiler or other equipment: part of the biomethane could be also sold to a refueling station for methane powered cars.

In order to reach medium pressures, a multi-stage compressor is necessary and its cost is around 4,000 €/kW. Medium-pressure storage (at 7 bar) specific cost is higher due to compression for biomethane too: investment is equal to 90.55 $ in Wattanasilp's paper [37].

2.6 Steam methane reforming (SMR)

Hydrogen production is composed of two reactions, a reforming reaction in which water in steam phase is added to react with hydrocarbons and produces hydrogen and carbon monoxide, and a shift reaction that conveys carbon monoxide oxidation and hydrogen formation. Hydrogen reforming process is very well known and, nowadays, the cheapest and most diffused way of producing it: according to Franchi's paper, currently 48% of hydrogen production comes from Steam Methane Reforming, 30% from petroleum fraction and 18% from coal gasification [38].

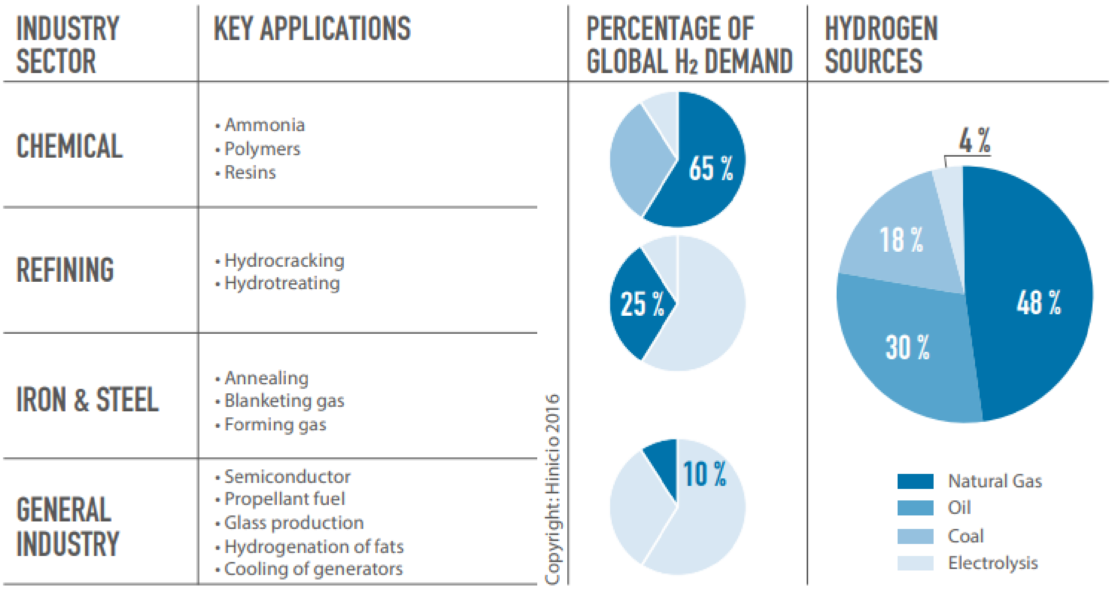


Figure 15 - Hydrogen sources and applications from IRENA based on FCH JU (2016) [39]

IRENA studies on hydrogen confirm such values in the Technology Outlook for Energy Transition [39].

The process starts with the steam methane reforming reaction, a boiler transform water into steam and then, it mixes it with biogas or biomethane in the reforming reactor. There, thanks to high temperature (between 750 and 1450°C [40]) the reaction occurs: hydrogen and carbon monoxide are produced:

$$CH_4 + H_2O \rightarrow 3H_2 + CO$$

Equation 2

The reaction is endothermic and needs 251 kJ/mol$_{CH4}$ [24], it's temperature dependent and its equilibrium constant increases with temperature [22] as shown in figure 16.

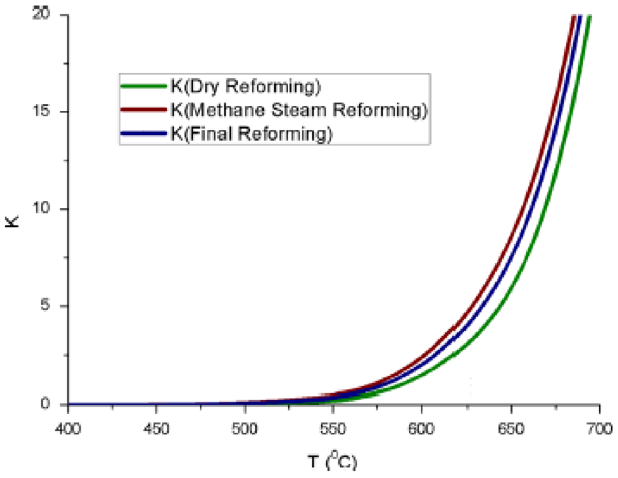


Figure 16 - Equilibrium constant in reactor as a function of temperature [22]

It means that the reforming reaction is enhanced with high temperature, but to reach higher temperatures, a higher amount of energy must be used. "Hydrogen production by biogas steam reforming" paper [22] reports that the best conditions are 1 atm of pressure and temperature between 700 and 900°C, while other literature sources [41] [42] reported best temperature conditions at 800-1000°C.

Steam-to-carbon (S/C) ratio can be defined to the amount of steam that must be used for hydrogen production: the ratio usually is 3:1 to obtain higher conversion levels. Conversion rate is directly proportional to S/C ratio and inversely proportional to pressure, as shows figure 17:

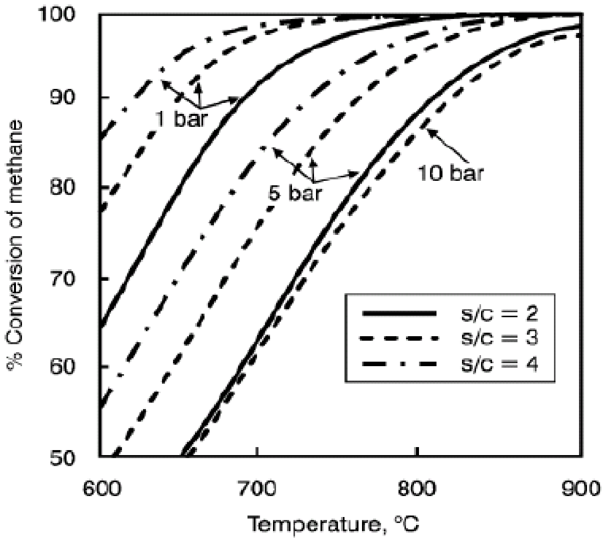


Figure 17 - Equilibrium conversion of steam reforming of methane against temperature, pressure and steam/carbon ration [43]

2.7 Dry reforming

In the first configuration purified biogas is directly sent to the reforming reactor. While CH_4 is involved in steam methane reforming reaction, CO_2 is implicated in dry reforming reaction: it occurs in the same reactor and hydrogen is produced:

$$CO_2 + CH_4 \rightarrow 2H_2 + 2CO$$

Equation 3

In literature [44], dry reforming ideal thermodynamics conditions are reported as: temperatures between 640 and 700°C, but with higher temperatures side reactions are minimized. With respect to steam methane reforming, the higher carbon presence may produce serious carbon deposition problems and CH_4/CO_2 ratio should be optimally adjusted: carbon deposition tendency is especially observed with low O/C and H/C ratios. It is quite difficult to increase both the ratios since an increment of the first one, the second one is reduced, and vice versa. Carbon deposition reactions must be accurately limited:

$$CH_4 \rightarrow C + 2H_2$$

Equation 4

$$2CO \rightarrow C + CO_2$$

Equation 5

These reactions can be inhibited with temperatures higher than 700°C [44].

Another side reaction that may occur is the reversed water gas shift, where hydrogen is used to produce water: speculations say that dry reforming will be abandoned for future hydrogen production [45].

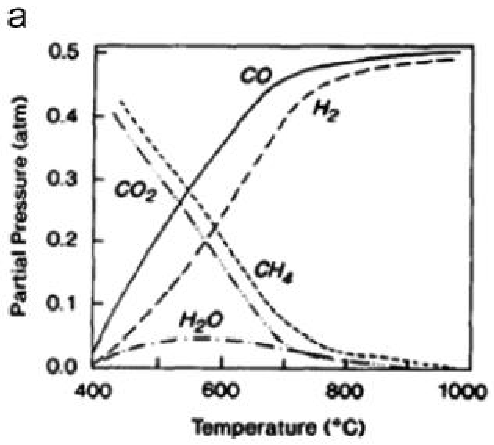


Figure 18 - Equilibrium gas composition fo CO_2/CH_4 (1:1) and 1 atm [44]

This reaction is not present in the second case because the reactor is fed by biomethane and carbon dioxide percentage is very small.

2.8 Water gas shift (WGS)

A shift reactor is needed to increase hydrogen yield, the reaction occurs thanks to steam water and carbon monoxide presence:

$$H_2O + CO \rightarrow H_2 + CO_2$$

Equation 6

It's an exothermic reaction ($\Delta H = -41.2\ kJ/mol_{CO}$) and takes place in a temperature range of 180-450°C. Due to low temperature in the shift reactor, an inter refrigeration heat exchanger is included: it cools down the flow and recovers thermal energy coming from the reformer reactor.

Multiple stages are usually considered for the WGS reactors: a high temperature shift (T=350-450°C) takes advantage of the high reaction rate and it is followed by a low temperature shift (T=180-250°C).

Catalyst is usually needed for water gas shift and the material used must include available oxygen vacancies, low carbon monoxide adsorption and activity in water dissociation [9]. Iron oxide catalyst is used for high temperature shift and copper, zinc and alumina oxide catalyst is used in low temperature shift [46].

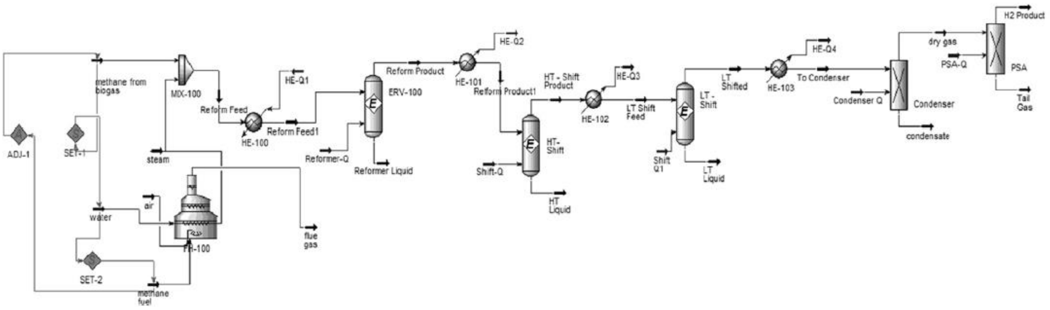


Fig. 1 – Plant of the hydrogen production process through the steam reforming of the biogas obtained in the fermentation of cassava wastewater.

Figure 19 - Detailed hydrogen production process scheme [37]

Steam reforming plant for hydrogen production (with 476 m^3/h H_2 output) is estimated to cost 3,142,897 $ [21].

2.9 Hydrogen purification

At the end of the shift reaction, the vapor flow is composed of 70-75% of H_2, 5-15% of CO, 10-15% of CO_2 and 5-20% of H_2O depending on the feedstock source and the production process [47]. Hydrogen separation is needed and many techniques are available:

- Pressure Swing Adsorption
- Membrane separation
- Metal hydride separation
- Cryogenic distillation

The assignment foresees a minimum hydrogen purity level of 99.975% to respect international standards ISO 14687 2:2012 for Proton Exchange Membranes (PEM) fuel cell uses [47].

2.9.1.1 Pressure Swing Adsorption

Separation relies on the different binding forces between the gases in the mixture and the adsorbent material. Hydrogen, and in general highly volatile components, are very difficult to be adsorbed, while bigger molecules, such as CO, CO_2, CH_4 and H_2O are adsorbed in the framework. Two or four columns compose the separation unit: when one is under adsorption operations, the second is in regeneration phase, and vice versa [48].

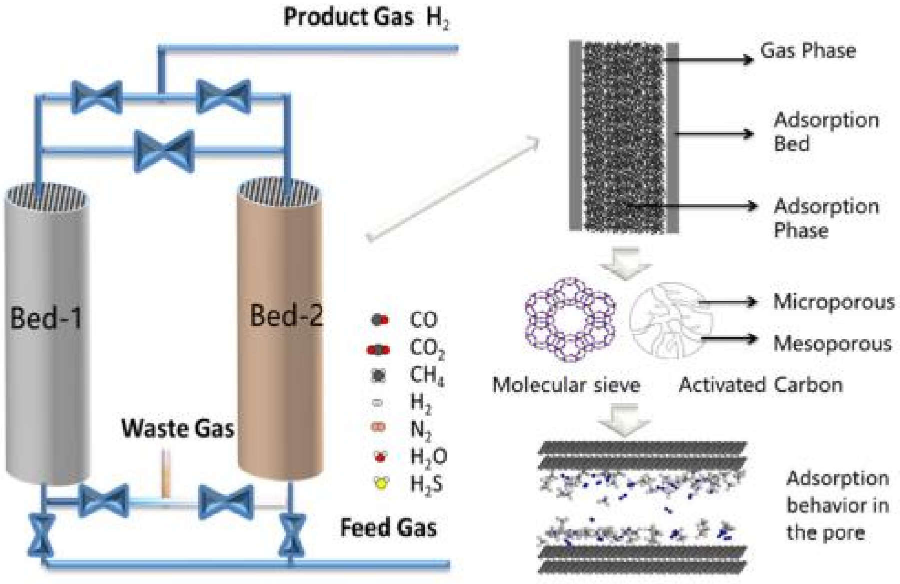


Figure 2. Flow scheme of the classical pressure swing adsorption (PSA) system.

Figure 20 - Flow scheme of the classical PSA system [49]

In literature, a lot of possible solutions are presented; by means of activated carbon or metal-organic framework (MOFs), different purity levels can be obtained, from 99.2% to

99.999% [48]. Interesting results had been obtained by Agueda in simulations with a new a new metal-organic framework named UTSA-16: 99.99-99.999% purity levels and 93-96% recovery rate had been obtained [50].

A very promising technology is the potassium-promoted layered double oxide described in Majlan paper: 99.9991% purity level can be reached and recovery rate accounts for 99.6% [51].

2.9.1.2 Membrane separation

Thanks to its compact structure, low energy consumption and flexible operation, membrane separation appears to be an encouraging technology: materials used are mainly polymers, metals and novel membranes (such as nanomaterials).

The working principle consists in the different permeation rate of the different gases through the membrane: similarly to the PSA, hydrogen is able to pass through, while bigger molecules remain blocked in the frame [48].

In literature [52], with metal membranes purification, hydrogen yields at 99.999% with CO content of 10 ppm.

2.9.1.3 Metal hydride separation

This technique is classified among the chemical separation methods and it consists in the use of H_2 storage to absorb/desorb hydrogen. With an increase in pressure and lower temperatures, H_2 is catalytically converted into H: impurities are blocked by metal particles, temperature increases, pressure lowers and impurity gases are discharged, lastly, hydrogen in crystal lattice is released [49].

Even though the presented technologies are very different among each other, purification methods cost is already accounted in production section: investment and maintenance cost are already comprehensive of that section.

2.10 Hydrogen compression and storage

Hydrogen must be stored to be transported and used later on in the refueling station (or eventually in the plant, replacing the hydrogen produced by REMOTE in case of maintenance): storage may be under liquid conditions, under gaseous phase with ad hoc compressors, or in metal hydrides (and low pressures).

Hydrogen liquefaction is reached at a temperature of 20.39 K: cryogenic storage must be very well insulated because even a very little amount of heat could easily produce hydrogen evaporation. Hydrogen molecule can be orto-hydrogen or para-hydrogen, for the first one proton have then same spin rotation, for the second one protons spin is antiparallel [53]. To liquify hydrogen, conversion from orto configuration to para is necessary: it naturally happens at low temperatures or catalytic technique is necessary. Most common processes used are Claude cycle or Linde process and are highly energy consuming, around 3.23 kWh/kg [54].

Hydrogen compression is a very popular storage technique because most of the transports (cars and buses) are built with 350-700 bar pressure tanks [55]. Typically, hydrogen is compressed at 200 bar and stored in cylindrical storages, hydrogen is a highly permeable molecule, for this reason ad-hoc tanks are produced with composite materials, such as fiberglass or carbon for structural support and polymeric materials for hydrogen permeability [56]. Compressors used for the process are categorized as turbocompressors if they give energy to the fluid by momentum variation, or as volumetric compressors if compression comes from volume reduction.

Metal hydride storage can be used to store hydrogen at low pressures (typically 10 bar) and it relies on hydrogen absorption in the metal lattice. Metals suitable for this work are heavy metals like titanium, vanadium and nickel: the problem of this storage technique is the weight and hydrogen mass only accounts for 2% of the total storage. Metal storage allows to reduce the storage volume, a very useful characteristic for transport solution, on the other hand, to store 4 kg of hydrogen (typical car storage dimension), the tank accounts for 200 kg: higher consumes are foreseen with such a solution [57].

Multistage compressor must be adopted to reach this pressure value, during compression, gas temperature increases and inter-refrigeration is necessary. Electric energy specific consumption for 4-stage compression is indicated in Tractebel report equal to 5 kWh/kg [55]. In the report it is explained that the compressor and auxiliary systems (such as the cooling circuit) have an overall efficiency of 50%; projections for future costs foresee the possibility to reduce compression cost of around 50% [55].

Marocco has identified in his paper the specific cost of hydrogen storage equivalent to 470 €/kg, operational costs accounts instead for 2% of the investment [58].

2.11 Hydrogen trailers

To transport hydrogen from the production center to the hydrogen refueling stations (HRS), some trailers are used. Hydrogen distribution has two main techniques: large bundle (a lot of small tanks grouped) or tube-trailers (bigger tanks of compressed hydrogen.

Two distribution procedures can be adopted: bump filling and even exchange. Bump filling consists in distributing hydrogen to stationary storages by means of pressure difference, multiple clients can be supplied with one tube-trailer and it's commonly done in Germany. Even exchange consist in empty-filled bundles or tubes swap: it simplifies supply to HRS and the client is only billed for the consumed gas (calculated with weight difference) [55].

Hydrogen battery

Hydrogen tube tuck/trailer

Figure 21 - Hydrogen batteries or tube trailers

Specific investment costs are reported in the document provided by Tractebel Engineering: tube trailers at 200 bar pressure cost 500€/kg of H_2 stored, while large bundles cost 470 €/kg [55].

2.12 Hydrogen refueling station (HRS)

Hydrogen refueling stations are the last piece of equipment necessary to provide hydrogen to the end-users. Currently, Europe can count on 140 HRSs and other 40 are under planned or under construction [59]. On the website H_2.live H_2 MOBILITY Deutschland GmbH & Co the list and position of the HRSs is present: it is easy to see that Germany and Belgium are now quite well covered already, instead large areas of Europe cannot be reached due to their complete absence [60]. The HRS will be positioned in La Aldea de San Nicolas de Tolentino, in the city center to provide refueling to Fuel Cell Electric Vehicles (FCEVs).

Nowadays passengers' cars predominantly have 700 bar storage technology, buses have 350 bar storage and forklifts operates at 250-350 bar (even if it is too early to define a trend).

HRSs are able to operate with refueling time that is compared to classic hydrocarbon fueling stations: the targeted refueling time is around 3-4 minutes and can be reached by means of hydrogen pre-cooling and infrared communication between FCEV and HRS. Currently, the main issue is the standardization metering accuracy of the dispensers. HRSs are under development and some studies must be done for the maximum amount of impurities that can be present in the fluid, to have a trade-off between fuel cell lifetime and hydrogen cost (linked to purification).

HRS investment cost is calculated by means of a cost function given by

$$CAPEX_{station} = A * \left(\frac{Q}{Q_{ref}}\right)^{0.66}$$

Equation 7

with A = 750 k€ (for 350 bar) or 1,500k€ (for 700 bar), Q_{ref} is the reference HRS capacity (200 kg/day) and Q is the selected capacity in kg/day. Operational costs are calculated instead as 4% of the CAPEX [55].

3 CASE STUDY DEFINITION

The considered case study is based in La Aldea de San Nicolás, a small village of Gran Canaria (Spain): it is part of the REMOTE European Project, which goal is to decarbonize remote areas and give renewable energy to alpine huts, companies and town based far from the national grid and any remote areas that, otherwise, should use hydrocarbon sources.

The chosen location has been selected among many other European candidates: presence of renewable energy sources, accessibility to the location, load profile appropriate for P2P solutions, business model and climate have been accurately analyzed for the site selection.

Near the municipality of El Furel, on Gran Canaria, a cattle farm by Grupo Capisa in expansion phase had been observed. It is located in the eastern part of the island, among three protected areas (Natural Reserve of Güi-Güi, Natural Park of Tamadaba and Natural Park of El Nublo): it is defined as an ecological island within a physical island. The whole municipality is a Protected Rural Soil, some constraints and limitations are imposed in the area for ecological and environmental protection, energy uses can be only allowed for agro-food facilities, electricity grid in the municipality cannot be upgraded and extended, and electricity can only come from renewable energy sources. Accessibility to the island is granted by an international port (4th in Spain in cargo ships traffic) and airport, both located in Las Palmas de Gran Canaria; from the capital, road connections allow to reach La Aldea de San Nicolás.

3.1 Business Case

As previously mentioned, Grupo Capisa owns a cattle farm for milking production; since a farm expansion, population has reached 1000 cattle. Current power consumption consists in 40 kW from the main grid of El Furel town, but additional 100-120 kW peaks are foreseen to be covered with the new electrical loads. Existing diesel generators are now used to cover power peaks but must be replaced by renewable energy source paired with batteries and other storage systems.

Additionally to the REMOTE system, a side project has been proposed: it consist in the conversion of cattle manure produced in the farm into hydrogen to power FCEVs that may populate the area in the next years and may need a refueling station and hydrogen energy sources. Once produced, hydrogen may also be used to supply the primary REMOTE project in case of RES systems maintenance or another expansion of the farm.

Two cases have been identified:

- Case 1: biogas is produced through anaerobic digestion of manure and it is, then, converted in hydrogen with SMR and WGS reactions
- Case 2: biogas is, again, produced through anaerobic digestion of manure and upgraded later on to biomethane. Lastly, it is converted in hydrogen with SMR and WGS

Ad-hoc purification steps are added in both cases, to purify biogas, biomethane or hydrogen.

3.2 Approach

The present work has the goal to evaluate the techno-economic feasibility of a manure-to-hydrogen conversion plant.

The very first step consists in the analysis of the source feedstock, quantitative analysis of manure production from cattle have been found in literature. Organic substrates are a very variable source: their composition is based on the animal (or vegetal) from which it comes from, on the nurture and, in general, the animal living conditions. Since La Capisa Group was not able to supply chemical composition of manure from their farm, many papers have been compared to obtain such values.

Then, manure-to-biogas conversion step is studied: no biogas conversion attempt has been reported from La Capisa Group, so a literature approach had been necessary. Studies about biogas production from manure have been analyzed to find information on chemical elements present in the mixture, before and after the anaerobic digestion, and especially their quantities.

Once the theorical biogas composition have been obtained, the different parts of the plant have been selected and their operation has been observed thanks to literature data. Biogas flowrate has been the first valuable quantity obtained that allowed to select and correctly dimension all the downstream components.

Different biogas purification systems (and upgrading, for second configuration only) have been considered and selected depending on the purity parameters and investment costs. Similar analysis has been conducted for storages a hydrogen section components.

Hydrogen production have been developed and simulated with Aspen Plus® software: the scheme of the plant had been developed to reproduce a Steam Methane Reforming (SMR) and a consequent Water Gas Shift (WGS) reaction. The first goal was to understand hydrogen production and the other species present in the mixture. The

Steam-to-Carbon (S/C) parameter had been defined in the software in order to be able to simulate different conditions: sensitivity analysis on steam inlet for SMR has been developed in order to find the best trade-off between hydrogen production and carbon deposition risk.

Successive hydrogen compression, storage, trailer and refueling station solutions had been compared to evaluate costs.

For every component in the plant costs (CAPEX and OPEX) have been defined and energy consumptions (thermal, cooling and electrical) have been calculated: such information were lately reported and evaluated in the economic analysis. Financial analysis reports the levelized cost of energy (LCOE) and hydrogen (LCOH) and allows to compare the two different configurations and estimate future cash flows for the investment.

4 TECHNO-ECONOMIC CHARACTERIZATION

The input data for the whole system design is the number cows (1000) present in the milking facility that leads to the manure quantity evaluation. La Capisa Group was not able to provide data for the amount of manure daily produced in the farm: literature paper from Politecnico di Torino have been very useful in the definition of the quantity [61]. The paper analyses biogas production from different sources (wastewater, municipal solid wastes, industry, crops and livestock effluents) and explains how yield is calculated.

Livestock effluents usually have low biogas yield and, due to its low density, a lot of thermal energy must be used [61]. A lot of biogas production plants mix crops with livestock effluents in the anaerobic digester to increase density and biogas yield. Depending on the type of animal raised, different quantities and composition of manure can be obtained, for cattle of 600 kg in weight, around 42.6 kg/day of effluents are produced [61]. For livestock, biogas production estimates are not based on the total effluent production, but more on the amount of Solid Volatile (SV) content of effluents: this value depends on the specific living conditions of animals in the farm (nurture, climatic conditions, etc...). The paper reports a daily SV production of 4.463 kg [61] and biogas yields between 200 and 350 Nm^3/ton_{SV}, depending on the source: an average value of 295 Nm^3/ton_{SV} has been taken for the calculations.

ANAEROBIC DIGESTER INPUT DATA		
Number of cows	1,000	cows
Manure production per cow	4.463	kg_{SV}/d
Total manure daily prod	4,463	kg_{SV}/d
	4.463	t_{SV}/d
Manure-to-biogas conversion	295	m^3/tSV
Biogas daily prod	1316.6	m^3/d
Biogas hourly prod	54.9	m^3/h

Table 7 - Anaerobic digester input and conversion data

With these information the raw biogas production accounts for 54.9 m^3/h, and daily production is equal to 1316.6 m^3.

Specific attention had been put on the dimension evaluation of the anaerobic digester: considering manure density of 350 kg/m^3 and a hydraulic retention time (HRT) of 30 days [26], [36], daily manure volume flow is estimated as 71.4 m^3/day and total volume contained in the digester is 2143 m^3. Theorical volume is then incremented of about 25%: this additional space will be occupied by the biogas produced during fermentation.

ANAEROBIC DIGESTER		
Time retention	30	days
Total manure daily prod	71.4	m^3/d
Manure volume in anaerobic dig	2,142	m^3
Total anaerobic digester's volume	2,857	m^3

Table 8 - Anaerobic digester dimensioning

Anaerobic digester operates at a temperature of 35°C [25], [32], [36].

Biogas yields depends on manure composition, digester thermodynamic conditions, digester and catalyst presence. Again, no information on possible biogas composition coming from La Capisa cattle have not been provided from the group, different values and range of values have been observed in various papers: random values (within the correct value range suggested by literature) has been chosen for biogas composition [61]–[64]. Literature reports molar fractions of element in biogas from animal wastes anaerobic digestion as:

- Methane: 45-70%
- Carbon dioxide: 30-50%
- Water vapor: 0.5-5%
- Nitrogen: < 4%
- Oxygen: < 1%
- H_2S: < 300 ppm

The following tables reports the supposed biogas composition:

Biogas	Molar fraction
CH_4	55%
CO_2	40.1%
H_2O_v	3.7%
N_2	0.5%
O_2	0.5%
H_2S	0.2%

Table 9 - Biogas composition supposed after the anaerobic digestion

Power consumption is divided between heat for maintaining temperature and electricity: different technologies for anaerobic digesters exist, they have different power consumptions, but an average value has been considered [63].

Once biogas is produced, it must be purifed: it contains water and pollutant species that must be removed not to have maintenance problems. Specific attention is put in H_2S remotion from the mixture because its presence (combined with condensed water) can produce sulfuric acid and corrodes tubes and plant components; it increases maintenance costs for reparation and it reduces the working time of the production plant. Compressors, tubes and gas storage tanks are the systems that may be affected by corrosion problems.

First of all, water removal is performed: at the exit of the anaerobic digester the molar percentage has been estimated equal to 3.7% (literature says that water content usually accounts for 5% [28]). In the present work, water is considered to be completely eliminated from the biogas flow thanks to high-efficiency physical absorption or chemical adsorption systems. Hydrogen sulfide remotion follows and different techniques can be adopted (membrane separation, adsorption on activated carbons, etc.): even though the technologies can be very different, the efficiency levels are very similar and quite high (around 95-98% [28]).

After purification stage, biogas flow is reduced to 53.4 m^3/h (so 1281.4 m^3/day), which consists in 59,2 kg/h. Cost have been found for such systems in literature [30], but for different plant dimensions: the value has been scaled to REMOTE project size, with a generic scaling factor of 0,7.

Type	Original dimensions	REMOTE		Scaling factor	Original cost	CAPEX
Biological by BiogasClean	200	54.87	Nm^3/h	0.7	108,100 €	43,715 €
Activated carbons by Siloxa	500	54.87	Nm^3/h	0.7	70,300 €	14,969 €
Caustic scrubber Sulfurex	500	54.87	Nm^3/h	0.7	127,000 €	27,043 €

Table 10 - Scaling of CAPEX for different biogas purification technologies

Similarly, operational costs for the three different technologies have been considered and evaluated with same scaling cost technique.

Type	Original dimensions	REMOTE		Scaling factor	Original cost	OPEX (€/y)
Biological by BiogasClean	200	54.87	Nm^3/h	0.7	11,100 €	4,488 €
Activated carbons by Siloxa	500	54.87	Nm^3/h	0.7	35,500 €	7,559 €
Caustic scrubber Sulfurex	500	54.87	Nm^3/h	0.7	64,595 €	13,754 €

Table 11 - Scaling of OPEX for different biogas purification technologies

Even though BiogasClean system has the highest CAPEX among the three technologies, its OPEX is reported as the lowest one: for this reason, it has been chosen for the plant. On the long term (10 years comparison is reported in figure), such choice reduces operational cost:

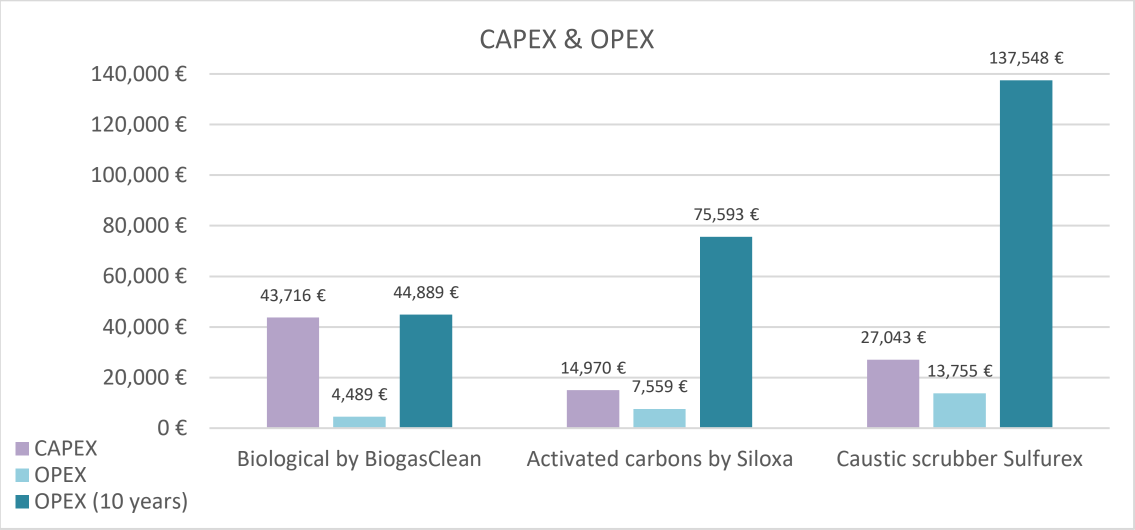


Figure 22 - Biogas purification CAPEX and OPEX comparison

Biological purification technique is chosen in order to reduce operational costs on the long-term period and obtain highly purified biogas.

In the same paper [30] annual electrical energy consumption for BiogasClean is equivalent to 9,000 kWh/year: electricity can be directly taken from primary section of REMOTE Project for free. The punctual power consumed is calculated with the running time of the machine:

$$Power = \frac{Annual\ consumption\ \left(\frac{kWh}{y}\right)}{Running\ time\ \left(\frac{h}{y}\right)} = 1.05\ kW$$

Equation 8

Annual consumption	9,000	kWh/y
Running time	8,500	h/y
Power (BiogasClean)	1.06	kW

Table 12 - Biogas purification power calculation

Data reported in “Biogas upgrading: Evaluation of methods for H_2S removal” from the Danish Technological Institute

The anaerobic digester and the biogas purification unit are common for both the configurations: the next sections are now divided between the first and second configuration. First configuration is analyzed in the beginning and lately, the second one.

4.1 Configuration 1: purified biogas

After biogas purification, a compressor system is paired to a storage: a deposit is placed at that point of the line in order to separate the biogas production from the hydrogen one. Such decision is taken to have separated lines that can be operated with different flowrates depending on the necessity of the moment, with low manure supply, biogas production is reduced while hydrogen plant can still convert the biogas into hydrogen and maintain the nominal production; same would happen with maintenance. In case hydrogen production section would be producing less than nominal flowrate, manure can be converted and biogas is stored in the accumulation system. Storages are used, also, to reduce peaks and lows; it is very useful because equipment can work with a defined flow rate and consumptions for start-ups/stops are reduced. This idea is proposed in the same way in the second configuration with biomethane.

For biogas storage, low and medium-pressure solution have been found in literature [36], [37], high-pressure storage (200 bar) suggested in papers [37] is evaluated as excessive and it would inefficiently increase investment and operational costs. Additionally, the hydrogen production section is operated at ambient pressure and high pressure is a useless consumption and waste.

Both for low and medium-pressure storage, a multi-stage compressor is necessary to overcome line pressure drops, and to reach 7 bar for medium pressure storage. The compressor has electricity consumption of 0.23 kWh/m^3 and with nominal flowrate, its instantaneous consumption is 12.3 kW [65].

COMPRESSOR DATA		
Electricity consumption	0.23	kWh/m^3
Cost of electricity	0.138	€/kWh
Compressor power	12.28	kW
Multi-stage specific cost	1,000	€/kW
Multi-stage compressor CAPEX	12,280 €	
Compressor OPEX	368.40 €	

Table 13 - Compressor consumption and investment calculatations

The holder is designed to store biogas for two days (48 hours) at nominal production rate, it can contain up to 2,843 kg.

Low-pressure storage is operated at ambient pressure, in this thermodynamical condition biogas density is equal to 1.11 kg/m^3 and the total volume is 2,562 m^3.

A first solution, from Petrollese and Cocco’s article has a cost of 40 €/m^3: the total investment would be 102,512 € but it is quite high, for this reason it is not taken into account.

In Wattanasilp’s paper on wastewater plant, he reports the specific cost of the storage equal to 6.94 $/m^3, so the investment would be around 15,117 € and O&M cost are 3% of the investment: 453 €/year.

In this report, medium-pressure storage is also analyzed and its cost is identified as 90.55 /m^3: biogas is stored at 7 bar with 7.77 kg/m^3 density, the total volume tank would be 366 m^3 and the capital cost is 28,179 €. OPEX is calculated as previously and it results 845 € [37]. Such solution is the most interesting one because it allows to have lower space occupation and higher energy for the biogas to flow in the next sections of the plant.

For the hydrogen production section, Aspen Plus® has been used to simulate realistic production flows and conversion rates for the section. First, the plant has been designed: input flows are biogas (which is the output of the storage) with a nominal capacity of 53.39 m^3/h and water flow that will be used in SMR reaction. A boiler is necessary to heat up water and transform it into steam, a mixer is then placed before the SMR reactor in order to have the inlet values (with one flow only) technically in a real plant no mixer will be present.

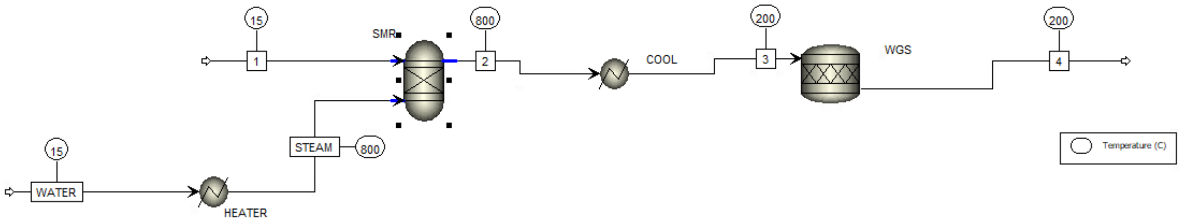


Figure 23 - H_2 production section designed on Aspen Plus® [31]

Biogas and steam enter the Steam Methane Reforming reactor: it has been modelled with a Gibbs’ reactor block with operating pressure at 1 bar and temperature at 800°C, according to previously presented data find in literature [22]. In the “Restricted Equilibrium” section, Steam and Dry Reforming reactions are defined. Molecular oxygen and nitrogen are set as inert elements. For Gibbs’ reactor block, a variable S/C has been defined in Fortran in order to easily set the Steam-to-Carbon ratio and perform sensitivity analysis. S/C ratio firstly set to 3 according to literature suggestions [66] and lately, it has been changed to understand its effect on production. Sensitivity analysis on S/C ratio testing values of 2, 2.5 and 3: the goal was to maintain a high conversion ratio without risking carbon deposition effect in the reactors.

Figure 24 contains the three different compositions at the output of the WGS reactor:

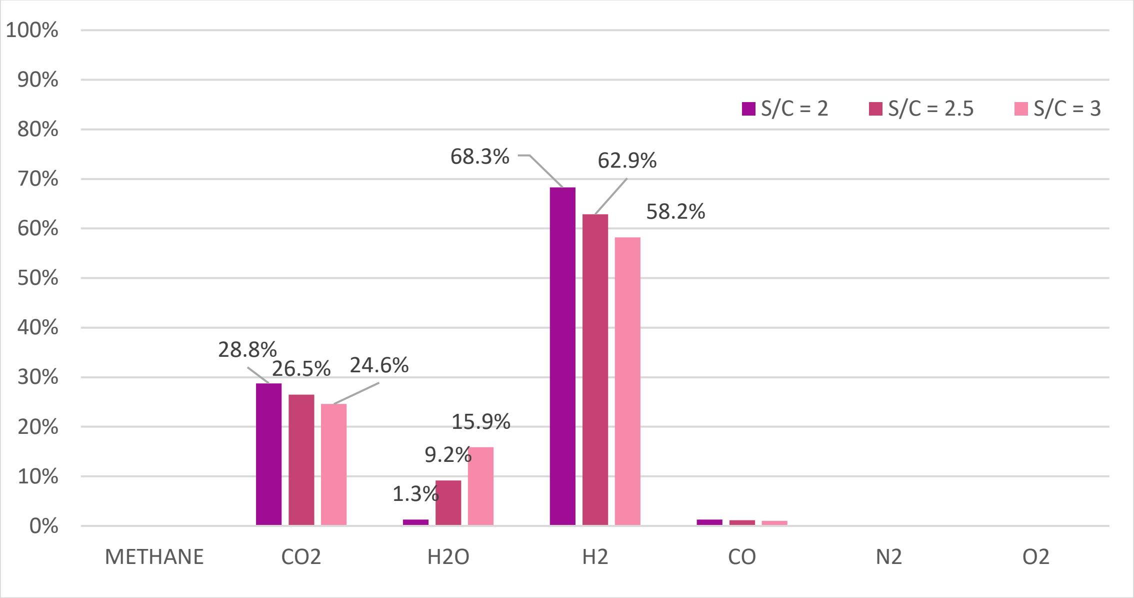


Figure 24 - Hydrogen conversion percentages with different S/C ratios

The reforming process for biogas comprehend both steam and dry reforming (SMR and DR, respectively with steam and carbon dioxide), the process efficiency has been set to 80%, according to Braga's paper [22]. In the same document, some analysis on the boiler for steam production are performed: boiler efficiency is set to 80% and 0.008 kg/h of biogas are necessary for every Nm^3 of hydrogen that enters the Gibbs' reactor [22].

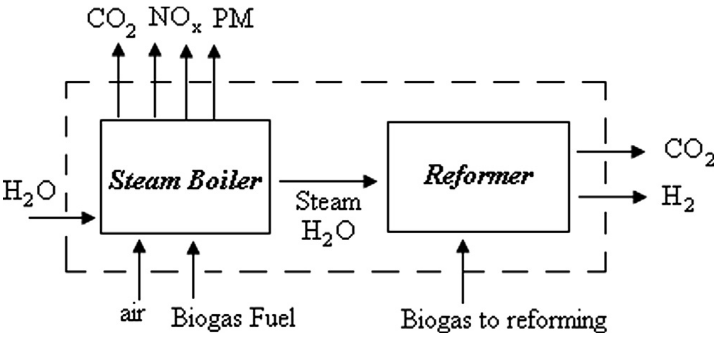


Figure 25 - Biogas boiler-reformer reactor coupling [22]

Biogas flow to the boilers directly comes from the storage previously analyzed: with Braga's specific consumption, the hourly biogas flow is evaluated in table 14.

Specific boiler biogas consumption	0.008	kg/h per Nm^3H_2
Total biogas consumed	1.506	kg/h
Percentage of biogas consumed	2.54	%

Table 14 - Biogas necessary for the boiler

Biogas consumed represents around 2.5% of the biogas produced in nominal conditions, the upstream biogas storage results useful also to power the boiler and being able to maintain nominal flow rate. Additionally, water pre-heating can reduce biogas consumption in the boiler.

Intercooling follows SMR to reduce temperature to 200°C, such heat is at relatively high temperature and further analysis should be performed to use it in other processes: it could be sent, for example, in a heat exchanger at the inlet of the boiler to pre-heat water. In general, this heat flow will be available to power some processes in the farm or reduce equipment's impact.

Once the flow has been cooled down, water gas shift is performed: it is used a RStoic at 200°C and 1 bar, fractional conversion of the stoichiometric reaction is set to 0.95. Simulation provides conversion rates and output flow composition shown in table 15:

Hydrogen	Molar fraction
H_2	62.86%
CO_2	26.51%
H_2O_v	9.17%
N_2	0.15%
O_2	0.15%
CO	1.16%

Table 15 - Raw hydrogen production flow composition

Aspen Plus® has been very useful to calculate heat and cooling power consumption of hydrogen production section:

POWER CONSUMPTION		
Heater	64.15	kW
SMR	106.13	
Cooler	-42.91	
WGS	-18.69	

Table 16 - Thermal power consumption in hydrogen production section

Hydrogen mass flow in nominal conditions is, then, reported in the following table:

ASPEN OUTPUT DATA		
Hydrogen produced	9.57	kg/h
Daily Hydrogen produced	229.8	kg/d
Hydrogen volume production	188.3	Nm^3/h

Figure 26 - Hydrogen output flow

Economic analyses are based on the price obtained in Madeira's paper: 3,142,897 $ needs to be invested to build the hydrogen production section of the plant, the research team had assessed the cost by means of the cost function that follows (equation 9) [21]:

$$CAPEX_{H2\;production} = 400 * \left(\frac{m_{H2}}{750}\right)^{0,5304}$$

Equation 9

The values for REMOTE plant had been obtained scaling the cost with a generic factor 0.7. Gim's paper reports similar economic evaluations: the team assessed a different cost calculation technique, but the final scaled cost is very similar to Madeira's article [21].

Type	Original dimensions	REMOTE		Scaling factor	Original cost	CAPEX
Gim [67]	152	229.8	kg/d	0.7	1,028,500 €	1,373,511 €
					Original cost	OPEX €/y
					159,800 €	213,405 €
Source	Original dimensions	REMOTE		Scaling factor	Original cost	CAPEX
Madeira [21]	476	229.8	kg/d	0.7	2,671,462 €	1,604,511 €
					Original cost	OPEX €/y
					267,146 €	160,451 €

Table 17 - Hydrogen production section CAPEX and OPEX scaling

Both the articles analyzed the cost of the whole SMR plant for hydrogen production: the investment cost already comprehend SMR and WGS reactors, cooling system, boiler to heat-up water and final purification system. In Madeira's dissertation the chosen purification system is the PSA [21].

Hydrogen compression and storage is the next step: for that section, the "Study on Early Business Cases" by FCH JU provided a lot of information for the economic evaluation. The investment cost of the compressor accounts for 687,000 € for 20 kg/h; even though the REMOTE plant has half of the production, the CAPEX is scaled as 2/3 of the given price (because smaller technologies have higher specific costs with respect to bigger once, for the economy of scale) [55]. Compression specific consumption is 5 kWh/kg to reach 200 bar pressure and accounts for 18,136 €/year.

HYDROGEN COMPRESSION		
Hydrogen production	9.57	kg/h
Compression CAPEX	453,420	€
Specific consumption	5	kWh/kg
Daily electrical consumption	48	kW
Compression OPEX	18,136	€/y

Table 18 - Hydrogen compression calculation

Hydrogen storage is then designed to have 2 days durability, and its specific investment cost is 470 €/kg:

HYDROGEN STORAGE SYSTEM		
Hydrogen storage (2 day)	459.56	kg
H_2 storage specific cost	470	€/kg
Storage cost	215,993	€

Table 19 - Hydrogen storage calculations

Compression power consumption is calculated as the specific energy consumption per the hydrogen flow rate:

$$Power_{compr}[kW] = Mass\ flowrate\ \left[\frac{kg}{h}\right] * Specific\ power\ cons.\left[\frac{kWh}{kg}\right]$$

Equation 10

POWER CONSUMPTION		
Compressor	48	kW

Table 20 - Hydrogen compressor power consumption

Last step consists in hydrogen transport to the hydrogen refueling station (HRS) by means of trailers. To do it, hydrogen is stored at 200 bar: each tube trailer can contain up to 500 kg of hydrogen, so at least 2 two tube trailers are necessary. Trailers must be accurately managed: with 500 kg, each trailer contains the amount of hydrogen produced in 2 days (229 kg/day). Every two days there will be an exchange of the trailers: the one to be refilled leaves the HRS and comes back to the plant, and vice versa. Depending on the real daily production of the plant, trailers may remain 1 or 2 days more, until they're filled or emptied.

Each trailer costs 500 €/kg, so, the CAPEX is 500000 €. Operational and maintenance costs for such products are evaluated as 4% of the investment: 20000 €/year [55].

HYDROGEN TRAILERS		
Pressure	200	bar
H_2 trailer storage	500	kg
Number	2	trailers
Tube Trailer CAPEX	500	€/kg
CAPEX (total)	500,000	€
Trailer OPEX	4%	-
OPEX (total)	20,000	€/y

Table 21 - Hydrogen trailer cost evaluation

The HRS is located in the municipality of La Aldea de San Nicolas de Tolentino and provides hydrogen to all the FCEVs (cars, buses and forklifts): it supplies hydrogen at 700 bar pressure for cars and 350 bar for heavy vehicle, such as buses. Its investment cost has been evaluated with the cost function provided by FCH JU [55]:

$$CAPEX_{HRS} = A * \left(\frac{Q}{Q_{ref}}\right)^{0,5304}$$

Equation 11

with A = 1,500k€ (for 700 bar), Q_{ref} is the reference HRS capacity (200kg/day) and Q is the selected capacity in kg/day. Operational costs are calculated instead as 4% of the CAPEX [55]. Investment values are summarized below:

HYDROGEN REFUELING SYSTEM		
CAPEX	1,643,908	€
$OPEX_{station}$	65,756	€/y

Table 22 - HRS CAPEX and OPEX

The study reports a specific power consumption equal to 3 kWh/kg

POWER CONSUMPTION		
Hydrogen supplied	9.57	kg/h
Specific consumption	3	kWh/kg
Power consumption	28.7	kW

Table 23 - HRS power consumption

4.2 Configuration 2: biomethane

In the second configuration, one more step is added: biogas upgrading is used to separate carbon dioxide and methane, the outlet flow is a biomethane. Biomethane is known as renewable natural gas and it is a mixture of methane and a small portion of carbon dioxide [68]. The additional transformation is interesting from because biomethane has more uses possibilities. Thinking about the energy transition phase, such additional step can be useful: at the begging FCEVs will be in very low numbers, but there will still be a lot of natural gas vehicles. Natural gas could be sold to these vehicles while thermal engine cars are switched with FCEVs.

Since biogas flow is very dirty (CH_4 = 57.23% and CO_2 = 41.73%), mass flowrate is drastically reduced and the mass flowrate is 20.62 kg/h.

BIOMETHANE OUTPUT DATA		
Biomethane daily prod	738.90	m^3/d
Biomethane hourly prod	30.79	m^3/h
Pressure	1	bar
Biomethane density	0.670	kg/m^3
Biomethane mass flowrate	20.62	kg/h

Table 24 - Biomethane output data

Biogas upgrading simply consists in CO_2 collection from biogas flow: pressure water scrubbing (PWS), hot potassium carbonate (HPC) and pressure swing adsorption (PSA) are compared. The composition at the outlet of the upgrading system is summarized in table 25:

Biomethane	Molar fraction
CH_4	98%
CO_2	2%

Table 25 - Biomethane composition

BiogasClean, Siloxa and Sulfurex systems have been compared and since the original dimensions are much higher than the REMOTE plant, scaling factor is applied and new values are found.

Type	Original dimensions	REMOTE		Scaling factor	Original cost	CAPEX
PWS	500	54.87	Nm^3/h	0.7	1,323,500 €	281,824 €
HPC (chemical)	500	54.87	Nm^3/h	0.7	1,057,400 €	225,161 €
PSA	500	54.87	Nm^3/h	0.7	1,407,500 €	299,711 €

Table 26 - Table for CAPEX scaling evaluation of different biogas upgrading techniques

From Barbera's document, costs for REMOTE-sized upgrading systems are 250000 € [31], instead Beccucci's master thesis reported 380000 € investment for such section [32]. *OPEX* are evaluated in the article and accounts for 10-13% of the investment cost:

Type	Original dimensions	REMOTE		Scaling factor	Original cost	OPEX (€/y
PWS	500	54.87	Nm^3/h	0.7	179,581 €	38,239 €
HPC (chemical)	500	54.87	Nm^3/h	0.7	279,094 €	59,430 €
PSA	500	54.87	Nm^3/h	0.7	158,888 €	33,833 €

Table 27 - Table for OPEX scaling evaluation of different biogas upgrading techniques

Specific power consumption is obtained by the technical paper and shown in the following table:

Model	Electric Energy (kWh/m^3)	Heating Energy (kWh/m^3)
PWS	0.27	-
HPC (chemical)	0.20	0.67
PSA	0.24	-

Table 28 - Specific energy consumption model

With 54.86 m^3/h of biogas processed, the upgrading consumption is summarized in the figure below:

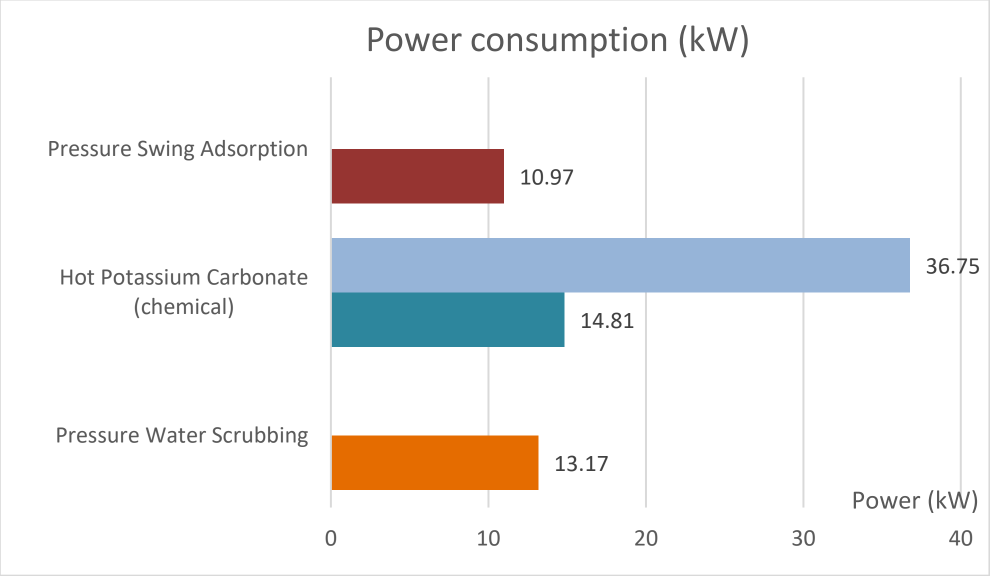


Figure 27 - Electrical and thermal power consumption

Biomethane is then stored (48 hours capacity) in order to provide a buffering phase for hydrogen production stage. The same concept to the previously analyzed biogas storage is applied to biomethane: similarly, the medium pressure storage system is preferred because it allows to reduce the necessary space without having high compression costs. Biomethane stored at 7 bar has density equal to 4.69 kg/m^3 and needs 211 m^3 volume tank:

MEDIUM PRESSURE STORAGE OUTPUT DATA		
Medium pressure storage	7	bar
Biomethane density	4.69	kg/m^3
Biomethane volume flowrate	4.4	m^3/h
Biomethane volume tank	211.1	m^3
Specific cost	90.55	\$/m^3
Total storage price	16,249	€

Figure 28 - Medium-pressure biomethane storage data

A second solution is explored: high-pressure can be useful if part of the methane is sold to CNG refueling stations and for space reduction. Biomethane can be compressed by means of a multistage compressor to 250 bar and stored in commercial natural gas high-pressure tanks: in James' document, 275 L 250 bar type IV CNG pressure vessels model is analyzed and manufacturers costs are provided [69]:

STORAGE OUTPUT DATA		
Storage time	48	h
Total mass stored	989.6	kg
Tank pressure	250	bar
Biomethane density	173.2	kg/m^3
Tank volume	5.71	m^3
Single tank internal volume	268	liters
Number of tanks	21	tanks

Table 29 - Biomethane high-pressure storage assumptions and analysis

In the review, the internal volume of each tank is reported to be 268 l, to cover the amount of biomethane produced at nominal flowrate by the plant for two days, 21 tanks are necessary. Each tank costs roughly 3,272 €:

STORAGE CAPEX	
Single tank price	3,272 €
Total tanks investment	69,748 €

Table 30 – Biomethane high-pressure storage CAPEX

Compression and cost storage costs comparison is reported in table 31:

Comp & Storage costs		
Type	CAPEX	OPEX €/y
High pressure storage (250 bar)	98,073 €	3,922 €
Medium pressure storage (7 bar)	44,573 €	1,782 €

Table 31 - Compression and storage cost comparison for configuration 2

Since 250 bar compressed biomethane has no purpose for the project, the choice fell on medium pressure storage that allowed to reduce costs and energy consumption.

Hydrogen production is now performed by the steam methane reforming only. Since CO_2 has been removed from the flowrate, the reactors are now scaled for a maximum daily biomethane flowrate of 30.8 m^3/h. The simulation process performed for biogas has been repeated with biomethane, instead. SMR reactor efficiency is assumed 80%, according to literature data [22].

The sensitivity analysis is, again, performed with values of 2, 2.5 and 3, to verify the best trade-off in conversion:

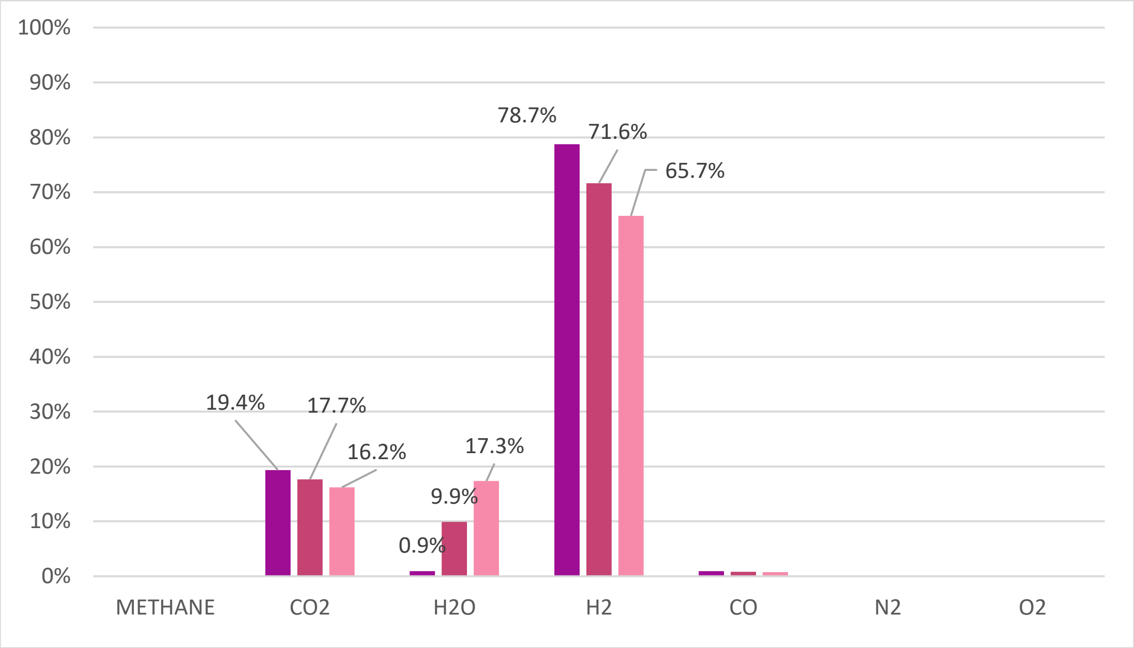


Figure 29 - Chemical composition of the output flow after Aspen Plus® simulation

Higher conversion rates are achieved with respect to biogas simulation because in this case, methane represents 98% of the mass at the inlet.

Finally, the hydrogen output is obtained:

ASPEN OUTPUT DATA		
Hydrogen produced	9.71	kg/h
Daily hydrogen produced	233	kg/d

Table 32 - Hydrogen output by Aspen Plus® simulation

Hydrogen is compressed stored and sent with trailers to the refueling station.

The amount of hydrogen is the slightly higher than the first configuration and investment for the last sections of the plant are the same:

HYDROGEN COMPRESSION		
Hydrogen production	9.71	kg/h
Compression CAPEX	453,420	€
Specific consumption	5	kWh/kg
Daily electrical consumption	49	kW
Compression OPEX	18,136	€/y

Table 33 - Hydrogen compression for biomethane configuration

Storage is included in the plant to have an additional buffer to smooth production peaks and maintain high hydrogen availability for the trailers even in case of low production:

HYDROGEN STORAGE SYSTEM		
Hydrogen storage (2 days)	466	kg
H_2 storage specific cost	470	€/kg
Storage cost	219,058	€

Table 34 - Hydrogen storage for second configuration

Two trailers are selected and for a total investment of 500,000€

The HRS is scaled on daily hydrogen production and the investment is calculated with the cost function given by the FCH JU document [55]:

HYDROGEN REFUELING SYSTEM		
CAPEX	1,659,264	€
$OPEX_{station}$	66,371	€/y

Table 35 - HRS CAPEX and OPEX for second configuration

5 RESULTS

5.1 Sensitivity analysis on S/C ratio

Sensitivity analysis is performed on steam-to-carbon ratio (S/C ratio) in Aspen Plus®. In literature, values higher than 3 are usually analyzed: in these simulations, values of 2, 2.5 and 3 have been tested in order to increase hydrogen yield in the reactors.

Molar percentages have been analyzed both after the SMR and WGS to understand the effect on both the reactors, both for biogas and biomethane configurations.

5.1.1 Configuration 1: purified biogas

After the reforming reactor, around 40% of hydrogen is produced, with a high amount of H_2O and CO.

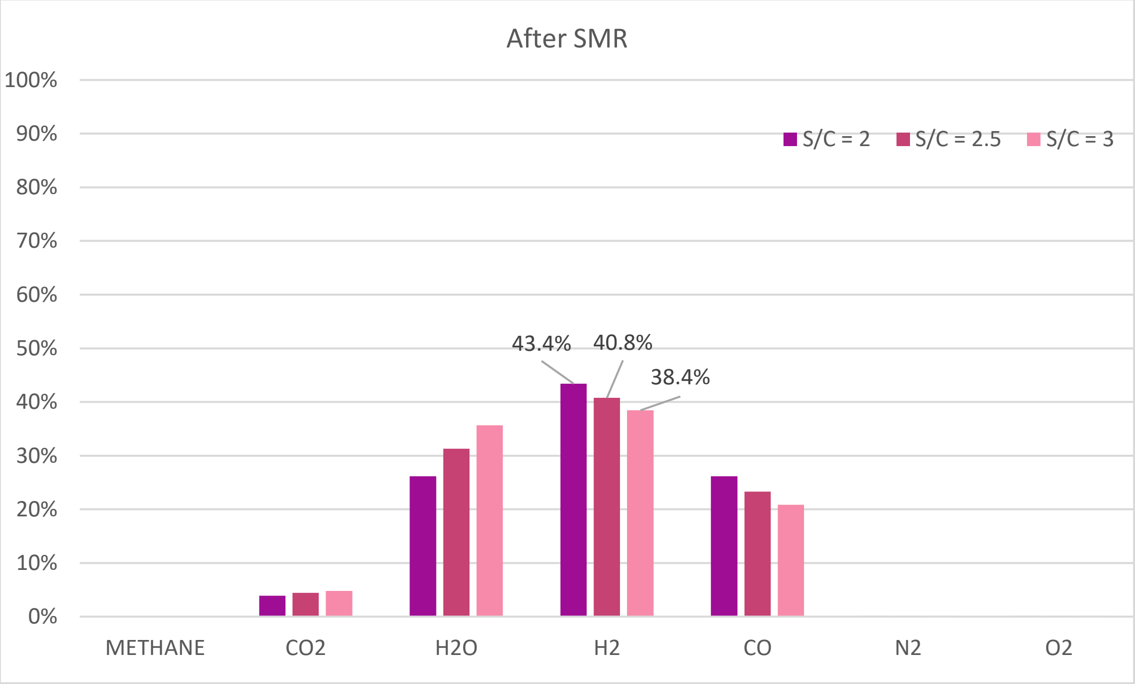


Figure 30 – Configuration 1: flow molar composition after SMR (sensitivity analysis)

Carbon monoxide is oxidized thanks to shift reactor and H_2 quantity is increased: the WGS effect is to extend H2 presence between 20-25% with respect to SMR output values.

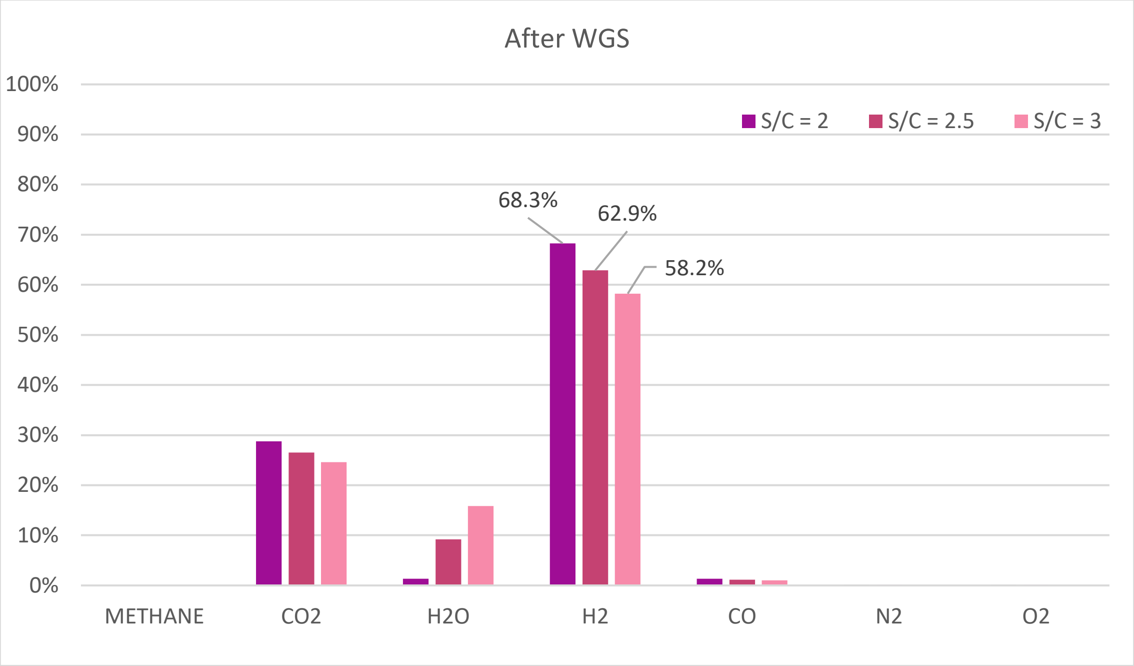


Figure 31 - Configuration 1: flow molar composition after WGS (sensitivity analysis)

Final yield is between 58.24% and 68.29%. By changing S/C ratio, 10% difference can be obtained. Also, water presence changes of about 13% between S/C = 2 or 3.

5.1.2 Configuration 2: biomethane

The effect of the S/C ratio variation on hydrogen production is roughly the same, but the starting point is different: the inlet flow is bio-methane (98% of methane) and its conversion is much easier. Just after SMR, values from 51.7% to 60.8% are reached:

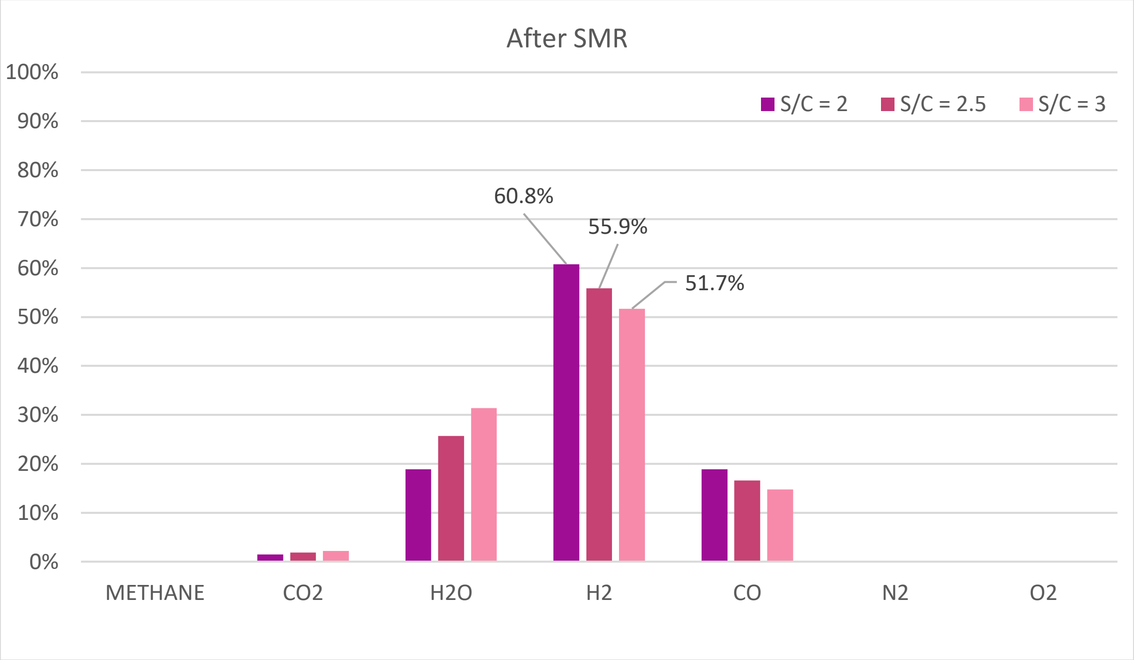


Figure 32 - Configuration 2: flow molar composition after SMR (sensitivity analysis)

S/C ratio regulation enhances hydrogen conversion both in SMR and WGS:

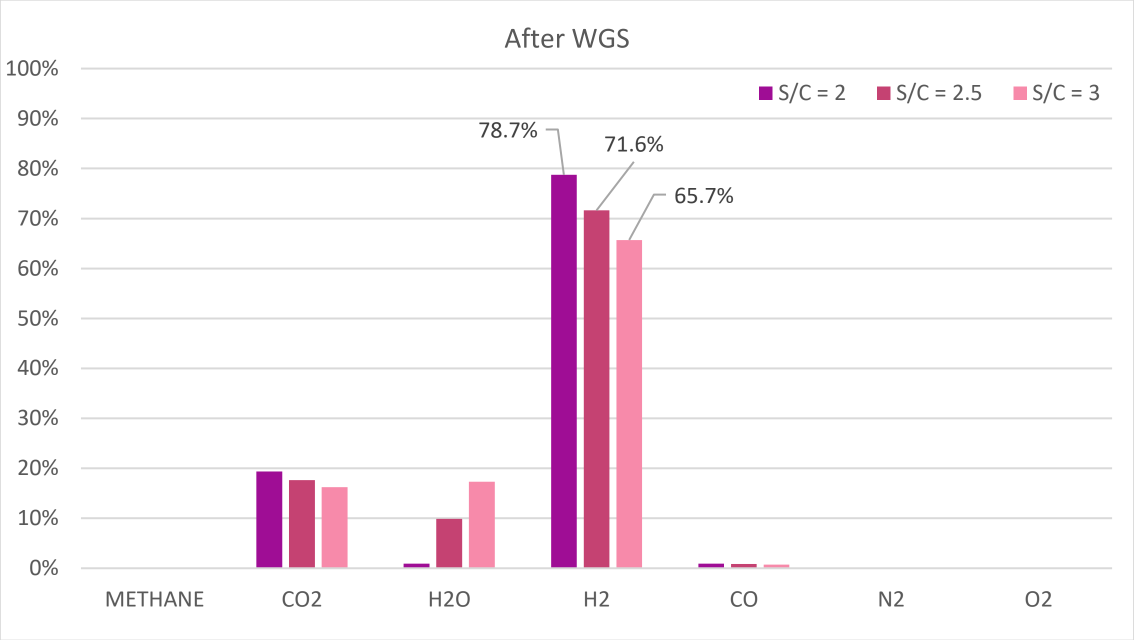


Figure 33 - Configuration 2: flow molar composition after WGS (sensitivity analysis)

Water gas shift reaction occurs and final hydrogen yield is between 65,7% and 78,7%

At the end of the simulations, the 2.5 S/C ratio is adopted in order to have higher conversion ratio than literature suggestions but maintaining enough steam in the mix to avoid carbon deposition. By choosing S/C = 2, higher probability of frequent maintenance and substitution costs are foreseen for such setting.

5.1.3 S/C ratio effects on power consumption

Even though molar percentages are higher in second configuration than the first one, the mass flow to which they are referred is lower and the final hydrogen mass flowrate is almost the same:

Hydrogen produced	Biogas	Biomethane
	9,57 kg/h	9,71 kg/h

Table 36 - Final hydrogen mass flowrates (after purification)

Heating and cooling power consumption can be calculated by means of Aspen Plus$^{®}$ and compared: with higher steam mass involved, higher heating and cooling energy is necessary in the processes.

Different plots have been made: first of all, a comparison is performed to compare equipment consumption and define a trend. The heater for steam production is the first equipment encountered and it shows a direct proportional consumption with the S/C ratio (and so, the amount of water that must be heated). SMR reactor has instead always the same energy consumption and do not depend on the S/C ratio. Cooler is again directly proportional to S/C ratio and WGS reactor has no dependance.

In general, heating and cooling directly depend on the mass flowrate:

$$W_{heat}, W_{cool} \propto \frac{S}{C} \propto m_{biogas}$$

Equation 12

5.1.3.1 First configuration

In yellow, the trend line shows the increase in the heater energy consumption with mass flowrate.

Reactors, instead, have a very stable consumption that do not show proportionality variation to S/C ratio because the thermodynamic conditions are the same in the three different simulations.

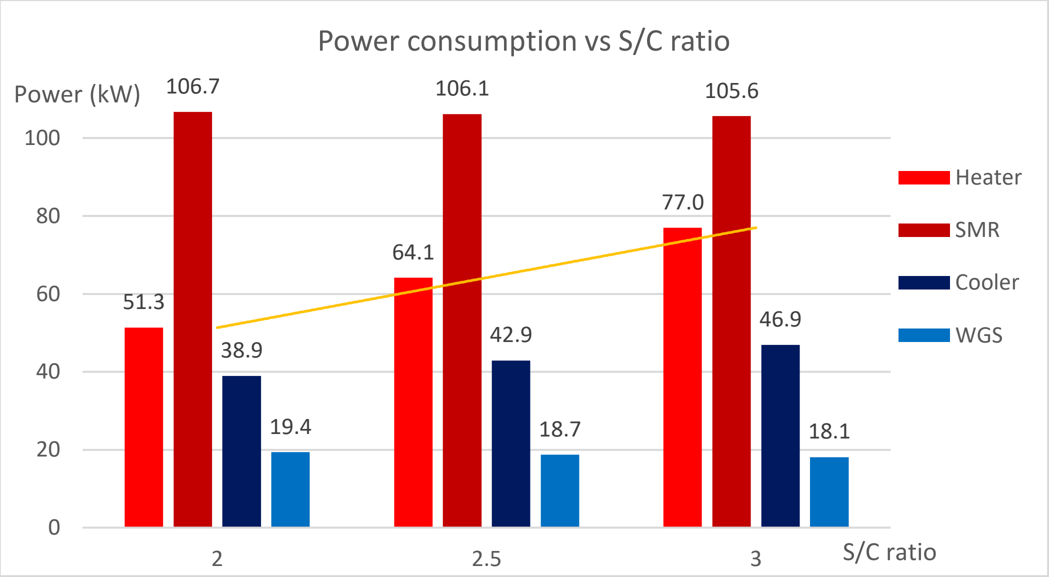


Figure 34 – Configuration 1: heating and power consumption of each equipment for different S/C ratios

A big picture graph is designed to show how much is the total power consumption increase: the increase previously described is present, but it is very low. Percentual variation to compare the three S/C ratios in heating and cooling energy is calculated:

$$Increase = \frac{W_{\frac{S}{C}=3} - W_{\frac{S}{C}=2}}{W_{\frac{S}{C}=2}}$$

Equation 13

Overall power and percentual increments are reported in table 37:

S/C	Heating (kW)	Cooling (kW)	Incr. Heating	Incr. Cooling
2	158,037	58,257	-	-
2,5	170,274	61,607	7,7%	5,8%
3	182,563	65,006	15,5%	11,6%

Table 37 – Configuration 1: overall heating and cooling power consumptions in hydrogen production section

Figure 35 reports a graphic representation of the percentual increments analyzed:

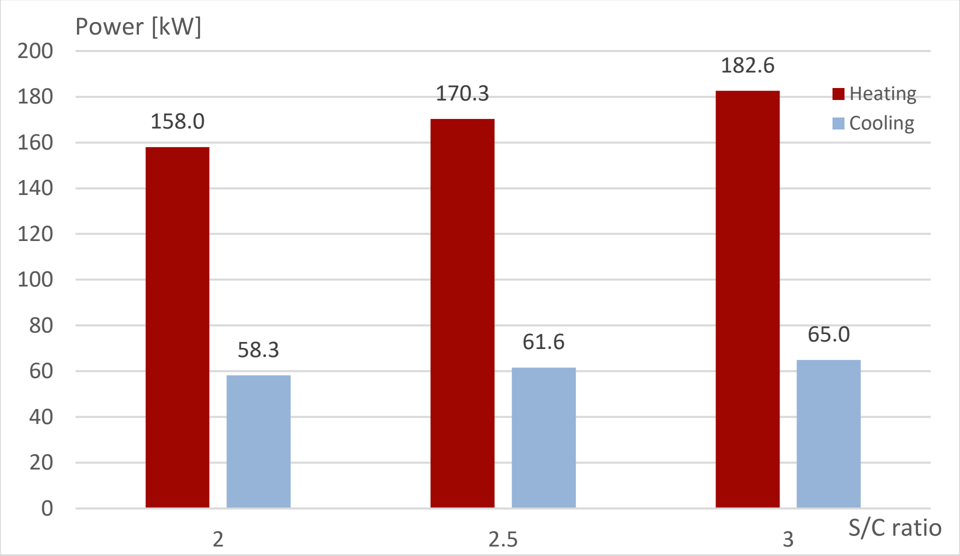


Figure 35 – Configuration 1: heating and power consumption for different S/C ratios

5.1.3.2 Second configuration

The same analysis is performed on second configuration (biomethane):

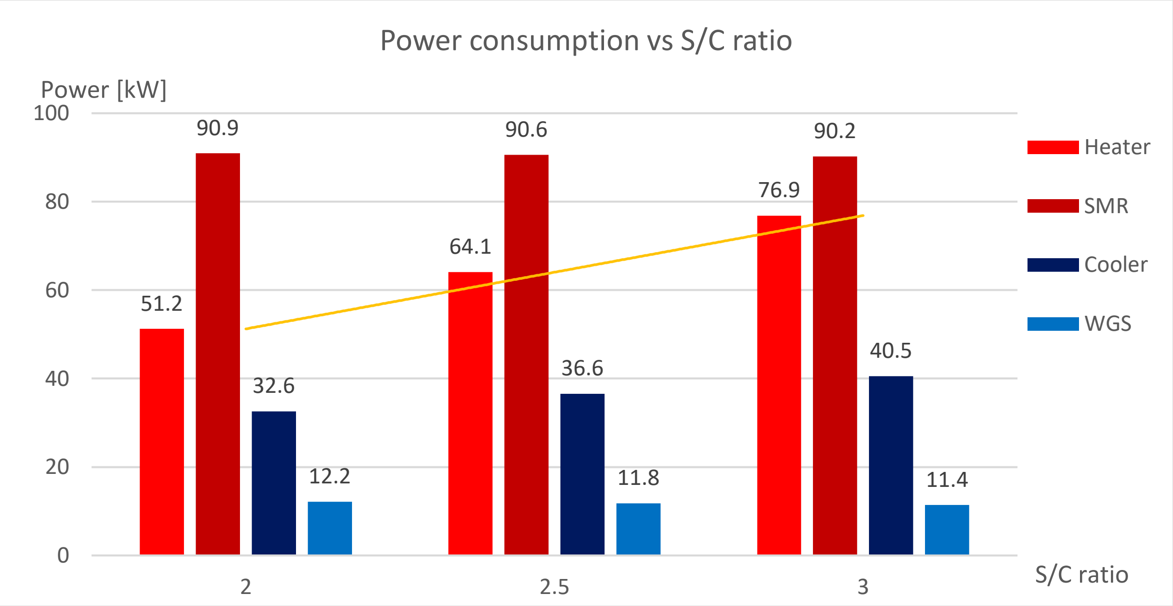


Figure 36 - Configuration 2: heating and power consumption of each equipment for different S/C ratios

Heater and cooler consumptions are the same as biogas-based plant, while SMR and WGS have a reduction in energy request due to the absence of CO_2 that must be heated in the flow.

In order to better understand the effects of S/C variation, a table with heating and cooling percentual increment is summarized.

Overall power and percentual increments are reported in table 38:

S/C	Heating (kW)	Cooling (kW)	Increment Heating	Increment Cooling
2	142.174	44.767	-	-
2.5	154.625	48.33	8.8%	8.0%
3	167.102	51.919	17.5%	16.0%

Table 38 Configuration 2: overall heating and cooling power consumptions in hydrogen production section

Figure 37 reports a graphic representation of the percentual increments analyzed:

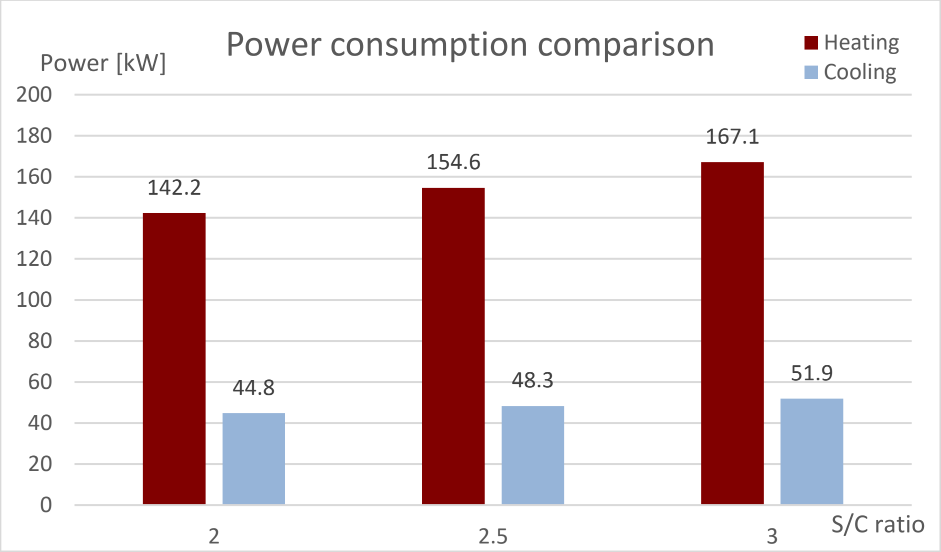


Figure 37 - Configuration 2: heating and power consumption for different S/C ratios

5.1.3.3 Observations on S/C = 2.5

From previous section, an S/C ratio of 2.5 is selected for plant design variable: such choice allows to have high hydrogen yield and low carbon deposition in the reactor.

The final power request for heating and cooling during hydrogen production process are summarized in figures 38 and 39:

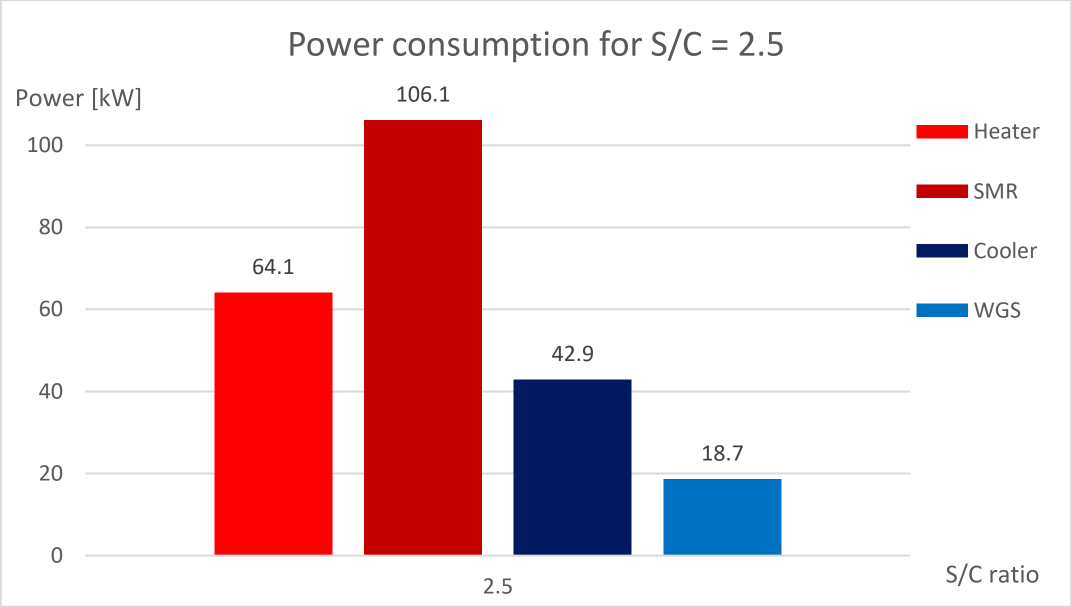


Figure 38 – Configuration 1: nominal power consumption divided among the different equipment

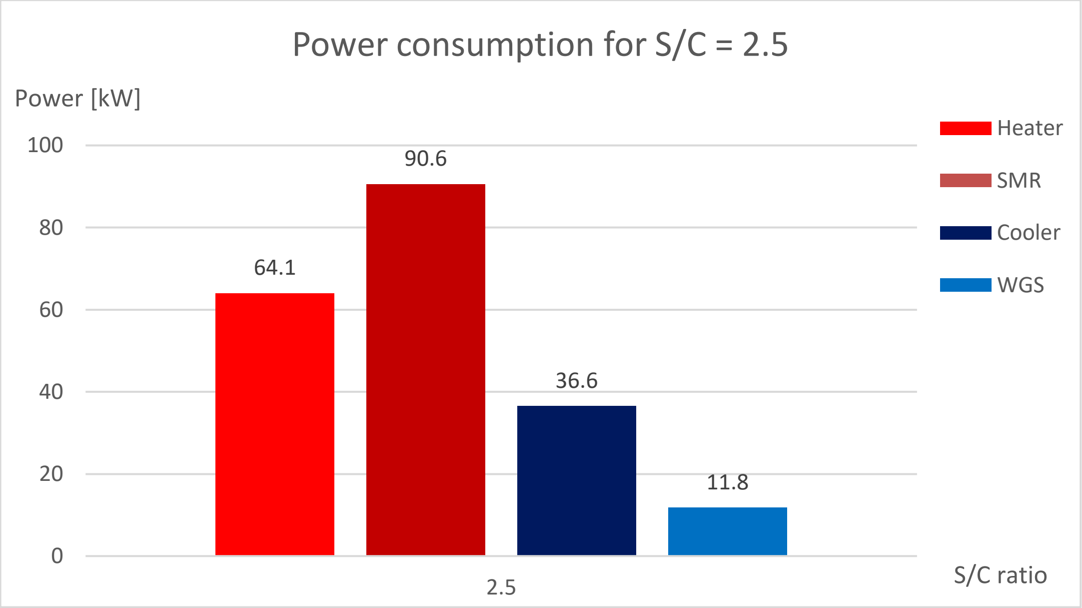


Figure 39 - Configuration 2: nominal power consumption divided among the different equipment

The heater needs the same energy in both cases, while all the other elements have different consumptions: the biomethane-based plant, in general, needs less energy due to the absence of carbon dioxide. Less matter is present in the flow and so, it will not be heated or cooled down.

Once S/C is selected equal to 2.5, the power consumption and economic analysis is developed.

5.1.4 Power consumption analysis

Power consumption has been performed to identify the energy duty that must be supplied to the plant to power all its processes: the main duty is brought by hydrogen formation process.

Anaerobic digester has the duty to maintain the manure-water mixture at 38°C for the fermentation process, according to Oehmichen's paper [63] 21 kW of heat and 3.18 kW of electric power are necessary for to preserve thermodynamic conditions:

POWER CONSUMPTION		
Process heat	0.107	MJ/MJ$_{biogas}$
LHV biogas	23	MJ$_{biogas}$/Nm^3
Biogas (energy)	6,191,312	MJ$_{biogas}$/y
Yearly Heat Duty	662,470	MJ/y
	184,021	kWh/y
Heat Power	21.01	kW
Specific Electric Power	0.0045	kWh/MJ$_{biogas}$
Yearly Electric Energy	27.860	kWh/y
Electric Power	3.18	kW

Table 39 - Anaerobic power needs calculations

Biological purification by BiogasClean is reported to have 9000 kWh/years for 8500 h/years: the electrical power duty is 1 kW.

Type		
Annual cons (BiogasClean)	9,000	kWh/y
Running time	8,500	h/y
Power (BiogasClean)	1.059	kW

Table 40 - Biological purification power consumption by BiogasClean

Biogas compression consumption for 7 bar storage is found in Gil-Carrera [65] and a 12.28 kW power compressor must be used to send 59.2 kg/h.

COMPRESSOR DATA		
Electricity consumption	0.23	kWh/m^3
Compressor power	12.28	kW

Table 41 - Compressor power consumption for 7 bar storage

Biogas upgrading has different technologies, PWS and PSA only need electricity while HPC needs both electrical power and heat. Consumptions are calculated from Barbera's analysis and then, plotted in figure 40:

	Electric power (kW)	Thermal power (kW)
Pressure Water Scrubbing	13.17	36.75
Hot Potassium Carbonate (chemical)	14.81	
Pressure Swing Adsorption	10.97	

Table 42 - Electrical power and heat consumption for biogas upgrading

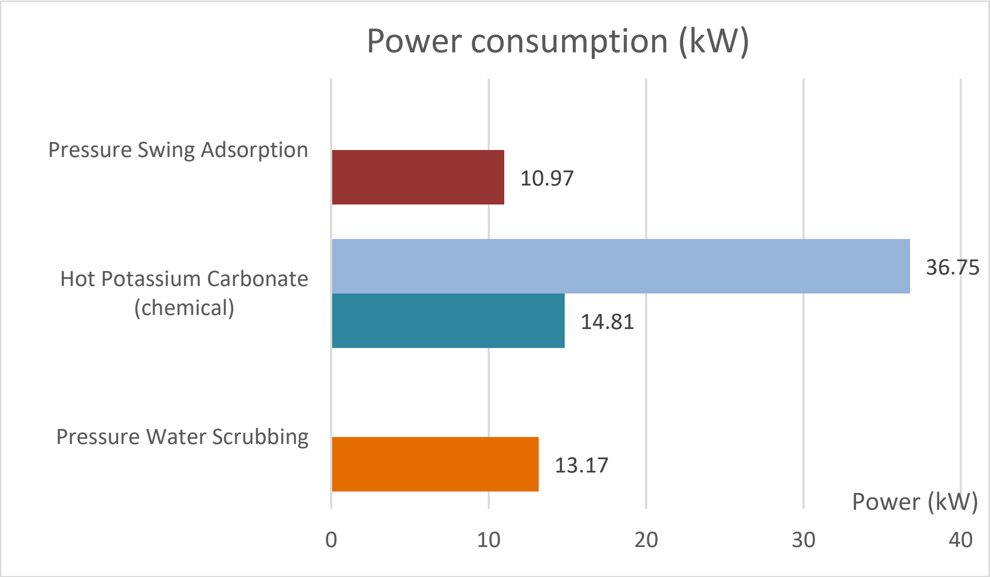


Figure 40 - Power consumption for biogas upgrading systems

Similarly to first configuration, biomethane is stored at 7 bar and the multistage compressor has 7.08 kW duty: it can process up to 20.62 kg of biomethane. Differently from the previous configuration, the mass flowrate that needs to be compressed is lower and a proportionality between the mass flowrate and electrical energy is defined:

$$W_{electr} \propto m_{flowrate}$$

Equation 14

There is no direct proportionality, but it's easy to see that with higher mass flowrate, the power consumption increases. Even though upgrading costs are withdrawn in first configuration, its absence implicates higher compression costs and power duties: CO_2 must be compressed and then heated or cooled down during hydrogen formation process.

The effects of upgrading presence or absence are shown on compression and storage section:

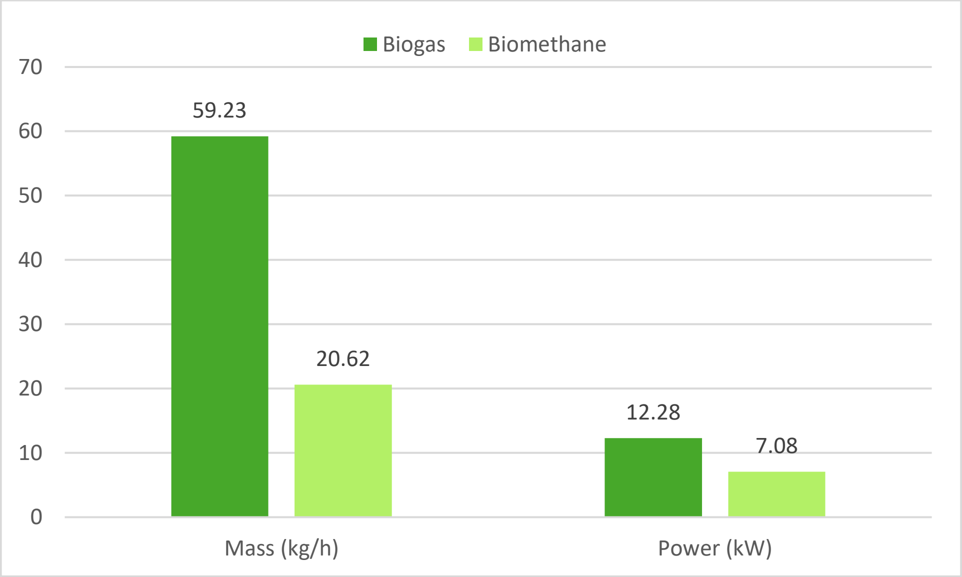
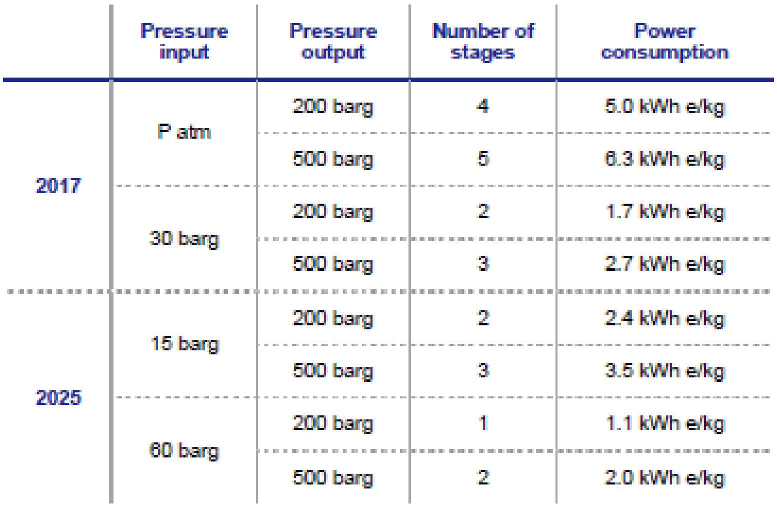


Figure 41 - Compression power depending on mass flowrate (effect of upgrading presence or absence)

Hydrogen production has been previously described, also from the energy duty point of view. Hydrogen compression is the next step and it is highly energy-intensive: in FCH JU report [55] the specific power consumption accounts for 5 kWh/kg for total 48 kW electric power.



	Pressure input	Pressure output	Number of stages	Power consumption
2017	P atm	200 barg	4	5.0 kWh e/kg
		500 barg	5	6.3 kWh e/kg
	30 barg	200 barg	2	1.7 kWh e/kg
		500 barg	3	2.7 kWh e/kg
2025	15 barg	200 barg	2	2.4 kWh e/kg
		500 barg	3	3.5 kWh e/kg
	60 barg	200 barg	1	1.1 kWh e/kg
		500 barg	2	2.0 kWh e/kg

Table 43 - Power consumption table from FCH JU report [55]

Lastly, the specific consumption of a HRS is 3 kWh/kg [55] and the total power is 28.7 kW_e (first configuration) and 29.1 kW_e (second configuration): the small difference is due to a very low hydrogen production variation.

Both in first and second configuration, the thermal, cooling and electrical energy consumption is summarized in a single table and total consumptions are found:

Configuration 1			
Component	Thermal Power (kW)	Cooling Power (kW)	Electrical Power (kW)
Anaerobic Digester	21.01	-	3.18
Biogas Purification	-	-	1.06
Biogas Compression & Storage	-	-	12.28
Hydrogen Production	170.28	61.6	-
Hydrogen Compression & Storage	-	-	48.00
Hydrogen Refueling Station	-	-	28.72
TOT	191.29	61.60	93.11

Table 44 - Configuration 1: power consumption for every equipment of the plant

Most of the heating power requested by the plant is necessary for the hydrogen production, divided between water heater and SMR reactor. The electrical power consumption is 26.9%, while heating accounts for 55.3% and cooling 17.8% only.

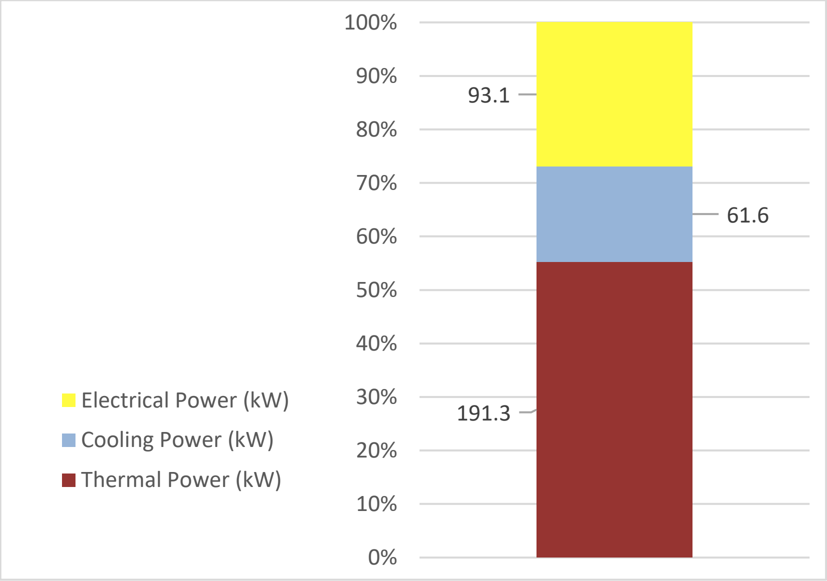


Table 45 – Configuration 1: biogas-based plant power consumption division

The specific energy consumption for electric, heating and cooling power have been calculated dividing the power for the hydrogen mass flowrate produced:

$$Specific\ energy = \frac{Power}{Mass\ florate_{H2}}$$

Equation 15

	Thermal Energy (kWh/kg)	Cooling Energy (kWh/kg)	Electrical Energy (kWh/kg)
Specific energy	19.98	6.43	9.73

Table 46 - Specific production energy

A breakdown of the three topics is reported in the following section.

Firstly, thermal request is observed. Anaerobic digester requires 11% of it and it needs low temperature heat because it only must maintain manure at 38°C for the fermentation process. The rest (89%) is asked in the hydrogen production section:

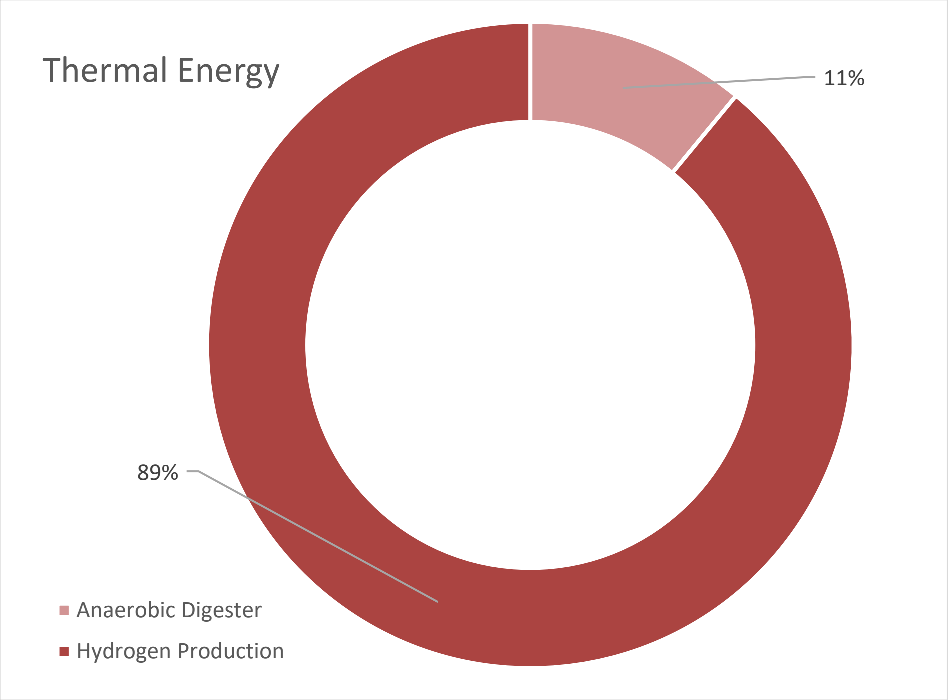


Figure 42 – Configuration 1: heating energy breakdown

Similarly, the breakdown analysis is done for the electrical power consumption: in this case every equipment consumes energy. A major role in the total consumption is related to hydrogen compression (more than 50%), by HRS (30%) and by biogas compression (more than 10%):

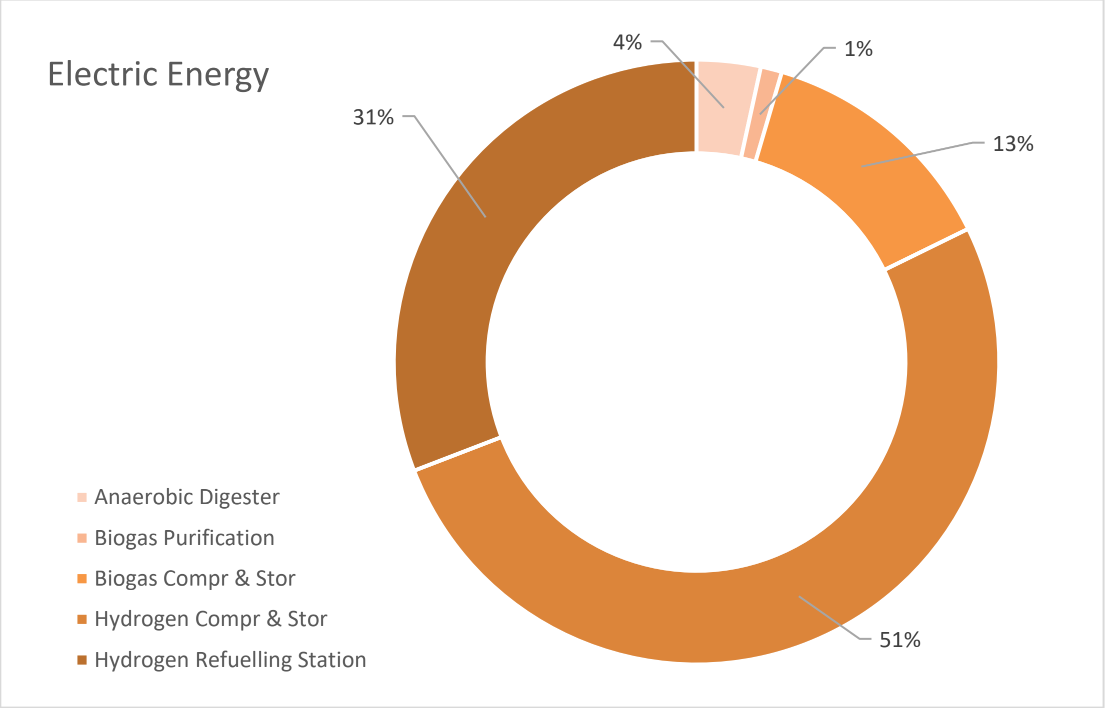


Figure 43 – Configuration 1: electrical energy breakdown

The same analysis is done for the second case:

Configuration 2			
Component	Thermal Power (kW)	Cooling Power (kW)	Electrical Power (kW)
Anaerobic Digester	21.01	-	3.18
Biogas Purification	-	-	1.06
Biogas Upgrading	36.75	-	10.97
Biomethane Compression & Storage	-	-	7.08
Hydrogen Production	154.62	-48.33	-
Hydrogen Compression & Storage	-	-	49.00
Hydrogen Refueling Station	-	-	29.13
TOT	212.38	48.33	99.97

Table 47 - Configuration 2: power consumption for every equipment of the plant

The electrical power consumption is 27.7%, while heating accounts for 58.9% and cooling 13.4% only.

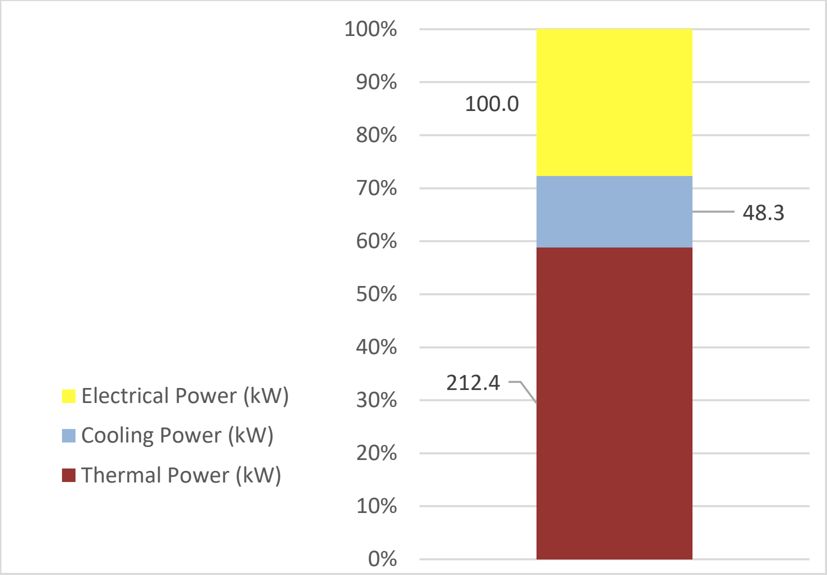


Figure 44 - Configuration 2: Biomethane based plant power consumption division

And specific energy for production is calculated too:

Specific energy	Thermal Energy (kWh/kg)	Cooling Energy (kWh/kg)	Electrical Energy (kWh/kg)
	21.87	4.98	10.30

Table 48 - Specific energy consumption

Thermal and electric power request breakdown is also reported in figures 45 and 46:

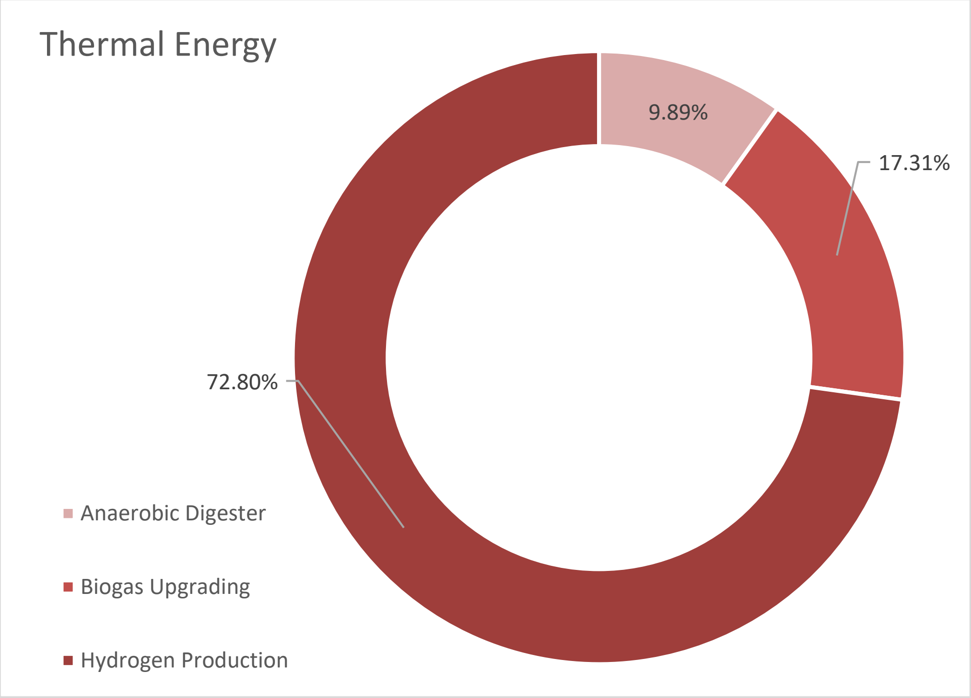


Figure 45 - Configuration 2: heating energy breakdown

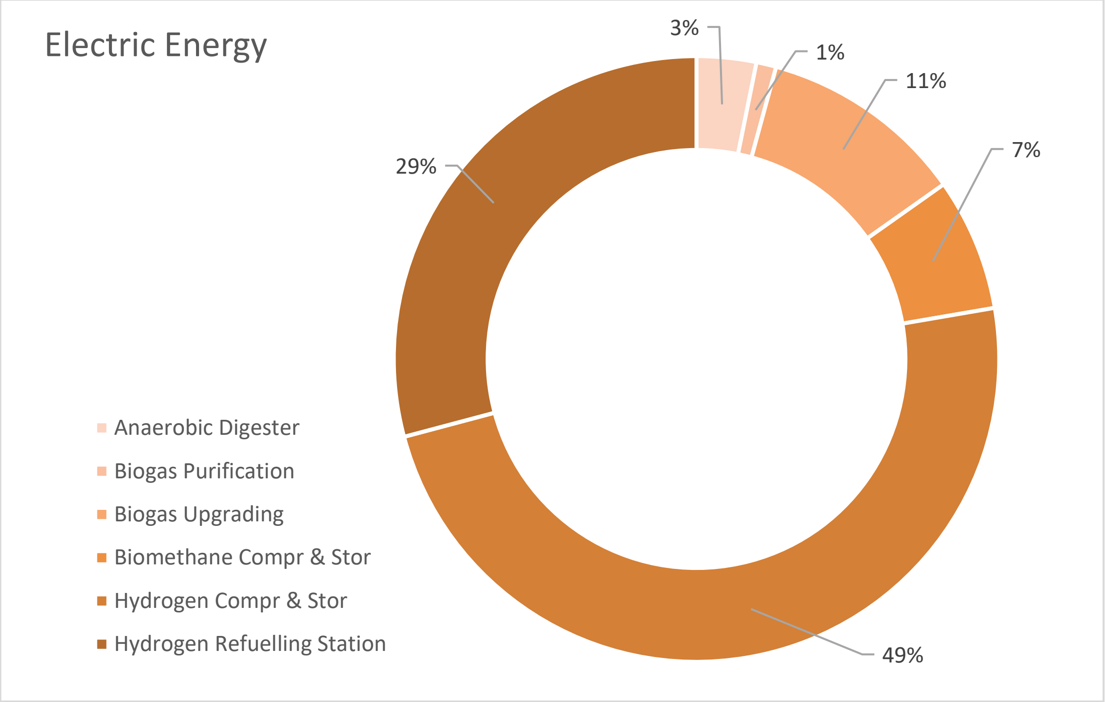


Figure 46 - Configuration 2: electric energy breakdown

Specific energy for H_2 production is shown in the following graph:

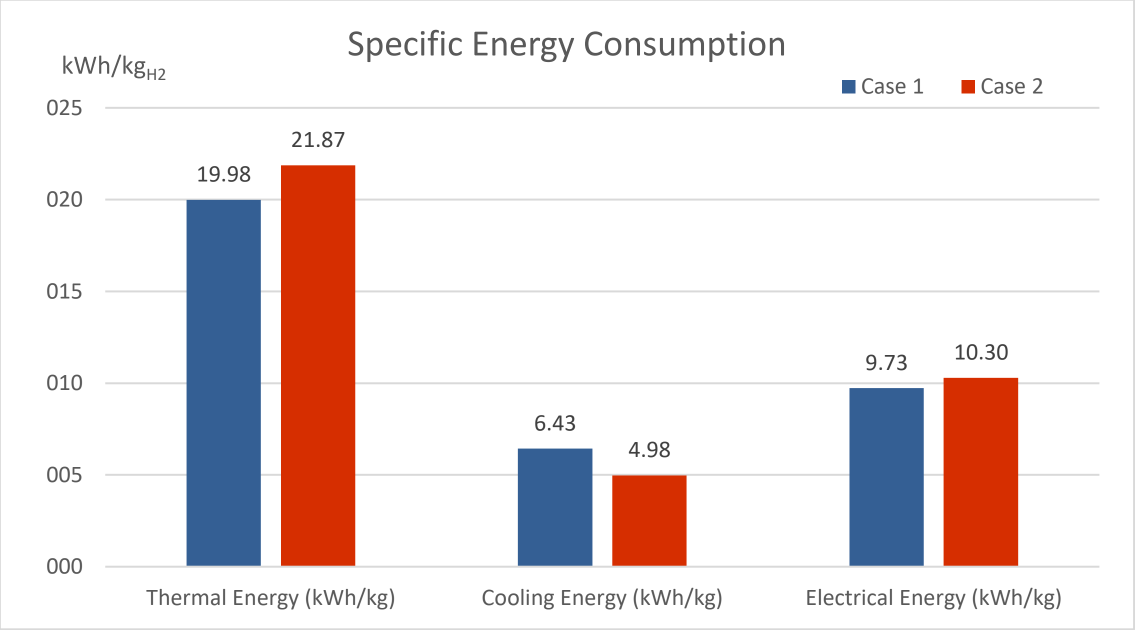


Figure 47 - Specific energy consumption comparison

In the figure an increase in thermal and electrical energy is observed, with a decrease of the cooling energy. In general, with the following values, a slightly higher cost in the hydrogen produced is expected in the second plant configuration: such effect is linked to biogas refining into biomethane, a highly refined by-product is obtained during the process and it has an economic impact.

5.2 Techno-economic analysis

Economic analysis has been carried out in order to understand the financial feasibility of the project. First of all, investment and operational costs have been evaluated for every piece of equipment: some technologies are also evaluated in order to provide an economic comparison and effects on the investment.

The following table summarizes the CAPEX and OPEX for every component, daily and annual hydrogen production:

Configuration 1		
Component	CAPEX	OPEX
Anaerobic Digester	571,428 €	74,250 €
Biogas Purification	43,715 €	4,488 €
Biogas Compression & Storage	40,459 €	1,213€
Hydrogen Production	1,604,511 €	160,451 €
Hydrogen Compression & Storage	669,413 €	18,136 €
Hydrogen Trailers	500,000 €	20,000 €
Hydrogen Refueling Station	1,608,460 €	64,338 €
TOT	5,037,987 €	342,878 €
Daily Production	229.78	kg/h
Annual H_2 production	83,869	kg
H_2 LHV	120	MJ/kg
Annual energy production	2.796	kWh
Selling price	9,38	€/kg
Revenues	786,698	€/y

Table 49 - Configuration 1: CAPEX and OPEX breakdown and production

The CAPEX and OPEX breakdowns have been reported in circle graphs to better picture the percentage division: the most influent elements resulted the HRS (32%), H_2 production (32%), H_2 compression & storage (13%), AD (11%) and H_2 trailers (10%). The total investment would more than 5 million € and the share is divided in this way:

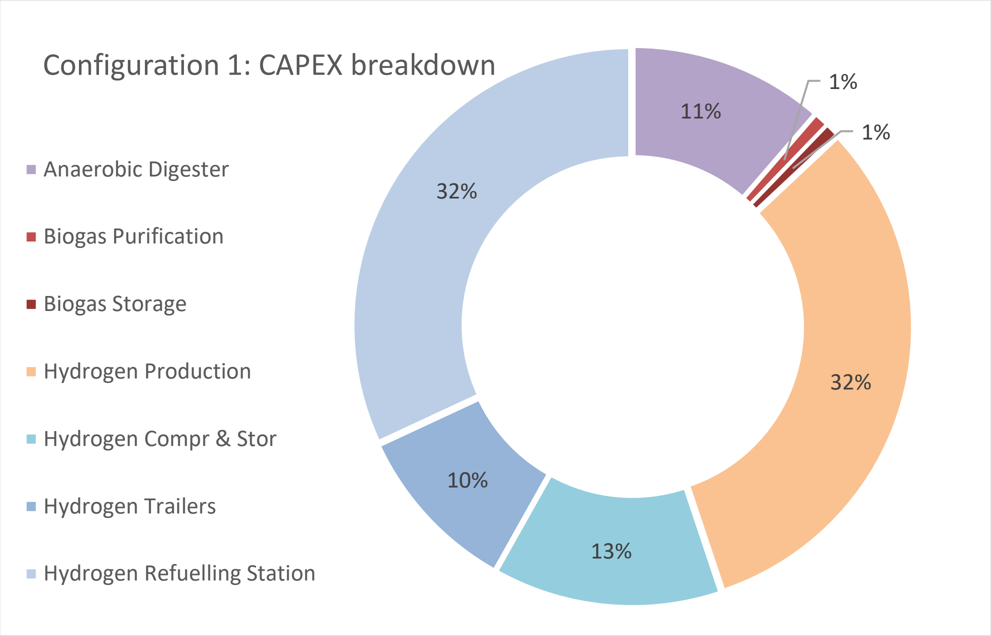


Table 50 - Configuration 1: CAPEX breakdown

In operational and maintenance costs H_2 production has the highest share (47%), then follow AD (22%), HRS (19%) and H_2 compression and storage and trailers (around 5% both):

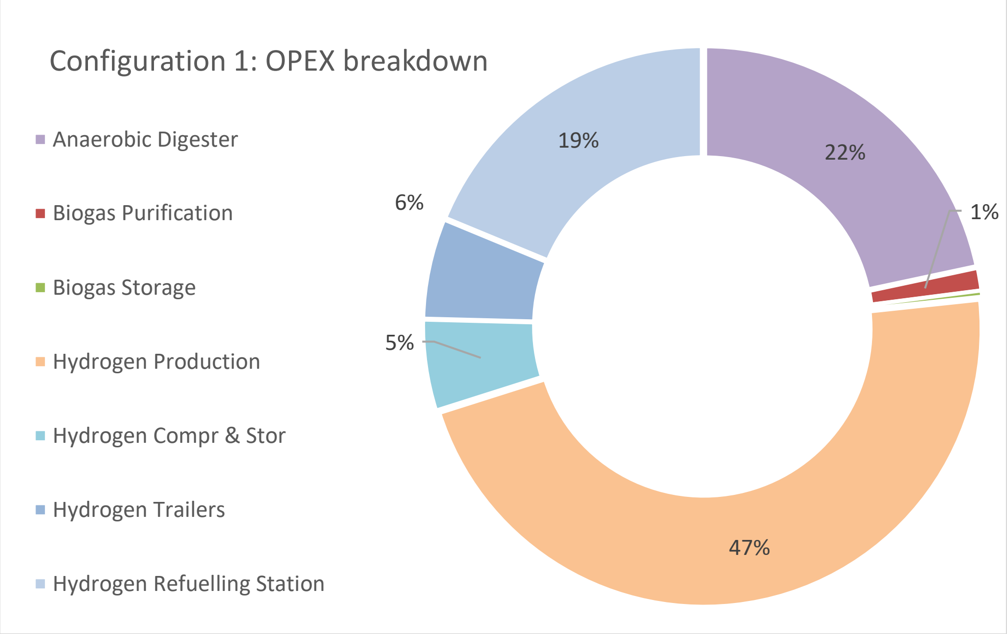


Table 51 - Configuration 1: OPEX breakdown

The high share of the H_2 production section is mainly due to the heater and SMR reactor thermal energy consumption. For all the equipment, the electrical energy consumed has been assumed to be free: the hydrogen production facility could be paired with REMOTE project that is able to provide renewable electric energy. Since REMOTE plant has been dimensioned for farms consumption only, the amount of PV panels, wind turbines and battery storages may be incremented to contain enough energy for the hydrogen production facility.

The following table summarizes all the component investment and operational and maintenance costs calculated for the second configuration, the annual hydrogen and energy production are listed below:

Configuration 2		
Component	CAPEX	OPEX
Anaerobic Digester	571,428 €	74,250 €
Biogas Purification	43,715 €	4,488 €
Biogas Upgrading	225,161 €	46,059 €
Biomethane Compression & Storage	44,573 €	1,782 €
Hydrogen Production	1,604,511 €	160,451 €
Hydrogen Compression & Storage	672,477 €	18,136 €
Hydrogen Trailers	500,000 €	20,000 €
Hydrogen Refueling Station	1,659,264 €	66,370 €
TOT	5,321,132 €	391,539 €
Daily Production	233.04	kg/h
Annual H_2 production	85,059	kg
H_2 LHV	120	MJ/kg
Annual energy production	2,835	kWh
Selling price	9.38	€/kg
Revenues	797,859	€/y

Table 52 - Configuration 2: CAPEX and OPEX breakdown and production

Also, figures for second configuration have been reported. For a total investment of more than 5.6 million € the highest percentages of the total investment are again for HRS (29%) and H_2 production (27%), H_2 compression and storage, AD, H_2 trailers and biogas upgrading follows (in order 12%,10%, 9% and 7%). Minor share is necessary for biogas purification and biomethane compression and storage.

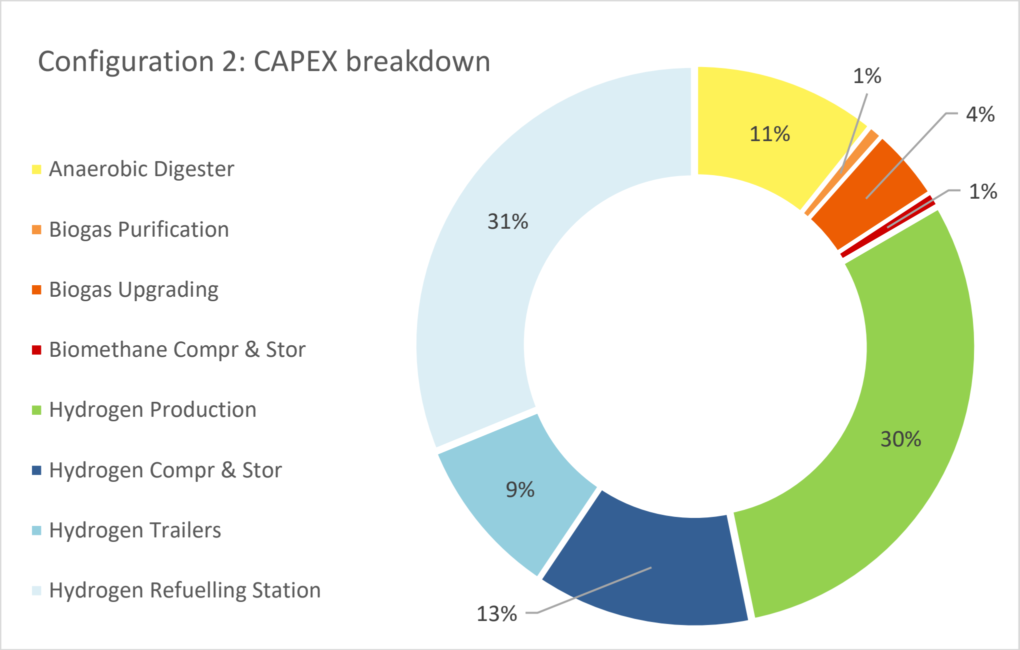


Figure 48 - Configuration 2: CAPEX breakdown

Similarly to configuration 1, H_2 production accounts for the highest share (40%) in the OPEX breakdown, after the AD (19%), the HRS (17%) and the biogas upgrading (12%):

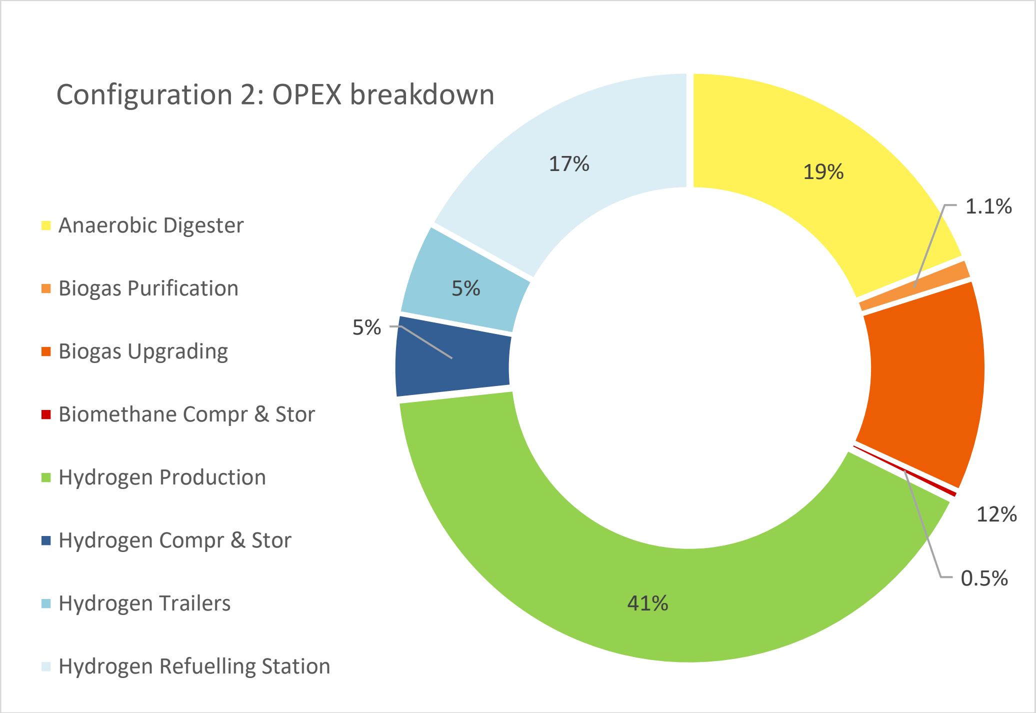


Figure 49 - Configuration 2: OPEX breakdown

The two configurations have been compared: the following figure reports the difference between CAPEX, OPEX and hydrogen daily production. In general, the second configuration has 10.63% higher CAPEX.

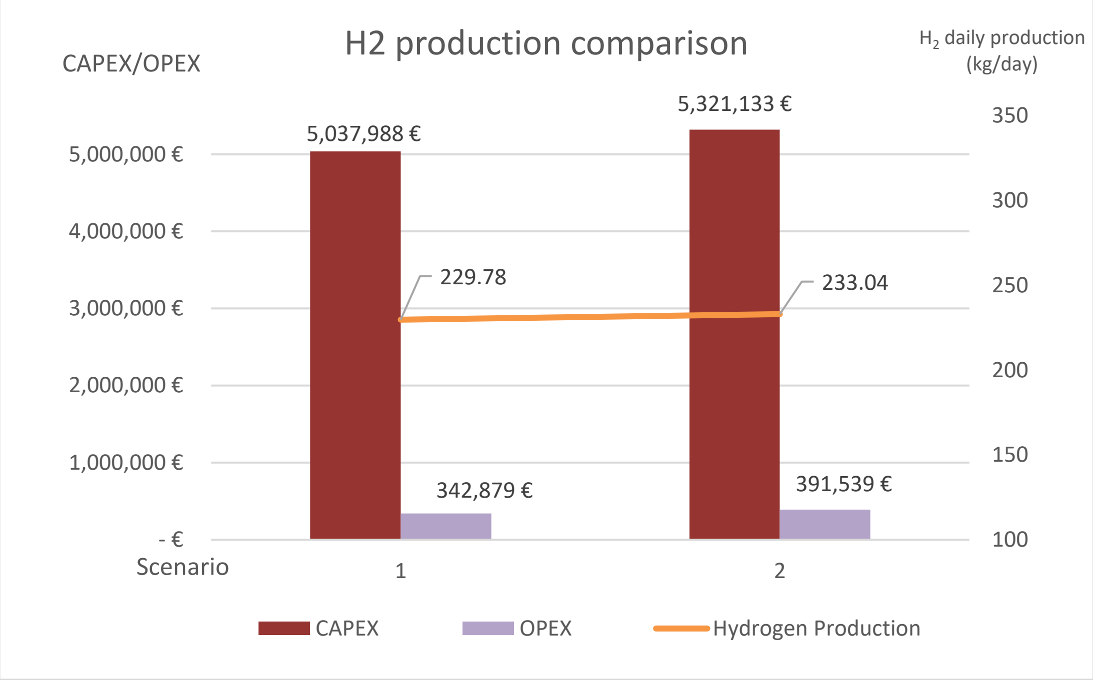


Figure 50 - Hydrogen production comparison between the two configurations

CAPEX, OPEX and hydrogen production have been compared by calculating the percentual variation between the two configurations: the investment increase has been 10.63%, while operational cost increased of about 13.92%. Instead, the hydrogen production increase has only been 1.40%: the mass flowrate increase has been lower than the financial percentage increase.

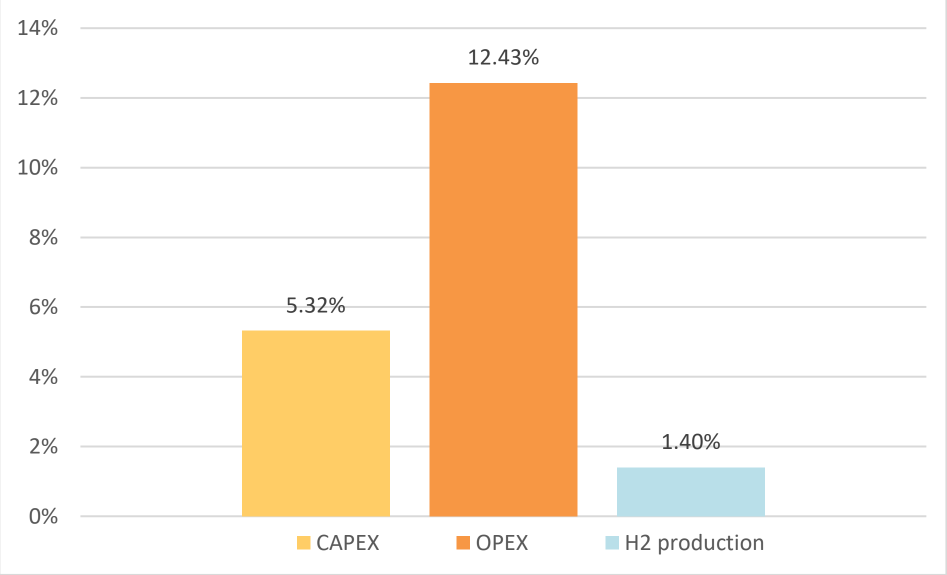


Figure 51 - Variation share between configuration 1 and 2

In general, second configuration construction is not justified by this increased flowrate amount: a good reason for this investment is the biomethane production as medium product. Biomethane could be useful, especially during the energy transition phase: biomethane can be used internally or sold to natural gas vehicles and applications, until a reasonable number of FCEVs will be present on the island.

Annual hydrogen production is calculated considering a 365 days production: thanks to buffer storages, during the maintenance period stored biogas/biomethane/hydrogen could still be supplied.

$$Annual\ H2\ production\ (kg) = Daily\ production\ \left(\frac{kg}{day}\right) * 365\ days$$

Equation 16

By means of hydrogen lower heating value ($LHVH_2$=120 MJ/kg), the annual mass production has been converted in energy production:

$$Annual\ energy\ production = Annual\ H2\ production * LHV_{H2}$$

Equation 17

Cash flow analysis has been evaluated: investment is carried out only in the first year and in the following years the operational and management costs have been considered only. Cash flow has been calculated as the difference between annual costs and revenues:

$$Cash\ flow = Cost - Revenues$$

Equation 18

Each year the cash flow must be actualized with respect to the first year (2021): for this purpose, the discount rate has been calculated as follows:

$$Discount\ Rate = (1 + WACC)^{-year}$$

Equation 19

And WACC (Weighted Average Cost of Capital) has been assumed to be 3%. This value represents the cost of capital for the investment and depends on how the project is financed: it especially depends on the cost of equity and cost of debt. Depending on how the project is funded, the effect is observed on actualized cash flow:

$$WACC = c_e * \frac{E}{E + D} + c_d * \frac{D}{E + D} * (1 - t)$$

Equation 20

The 3% WACC assumption is certainly a quite audacious choice: it represents a low investment risk.

Cash flow analysis is done for the first configuration and second configuration, considering three different selling prices: the tables are reported in appendix.

For the economic evaluation, Net Present Value (NPV) is a fundamental quantity that measures the profitability of an investment: it is defined as the sum of all the net cash flows evaluated during the whole lifetime of the project. It is calculated as follows:

$$NPV = -I + \sum_{t=1}^{n} \frac{(Net\ Cash\ Flow)_t}{(1+i)^t}$$

Equation 21

Positive NPV indicates a profitable investment and should be considered; a negative one is not interesting.

Internal Rate of Return (IRR) represents the discount rate *i* necessary to have null Net Present Value (NPV = 0), discounted cash flows equal to the investment:

$$-I + \sum_{t=1}^{n} \frac{(Net\ Cash\ Flow)_t}{(1+i)^t} = 0$$

Equation 22

With $IRR > discount\ rate$ the investment is considered a valuable investment, otherwise not.

Levelized cost of hydrogen and energy is obtained: it represents the minimum price at which hydrogen must be sold in order to return from the investment in 20 years:

$$LCOH = \frac{\sum_i Cost_i * d_i}{\sum_i Mass_i * d_i}$$

Equation 23

$$LCOE = \frac{\sum_i Cost_i * d_i}{\sum_i Energy_i * d_i}$$

Equation 24

Levelized costs have been calculated considering the whole supply chain (from the anaerobic digester to the hydrogen refueling station), LCOH and LCOE are evaluated at the outlet of the dispenser nozzle. With the given assumptions, the investment is foreseen to be economically viable and in line with current hydrogen cost trends. Sensitivity analysis is performed on cash flow analysis and three selling prices are proposed: 8.39 €/kg, 9.38 €/kg and 11.2 €/kg.

Cash flows are finally plotted in the graph and payback time is observed:

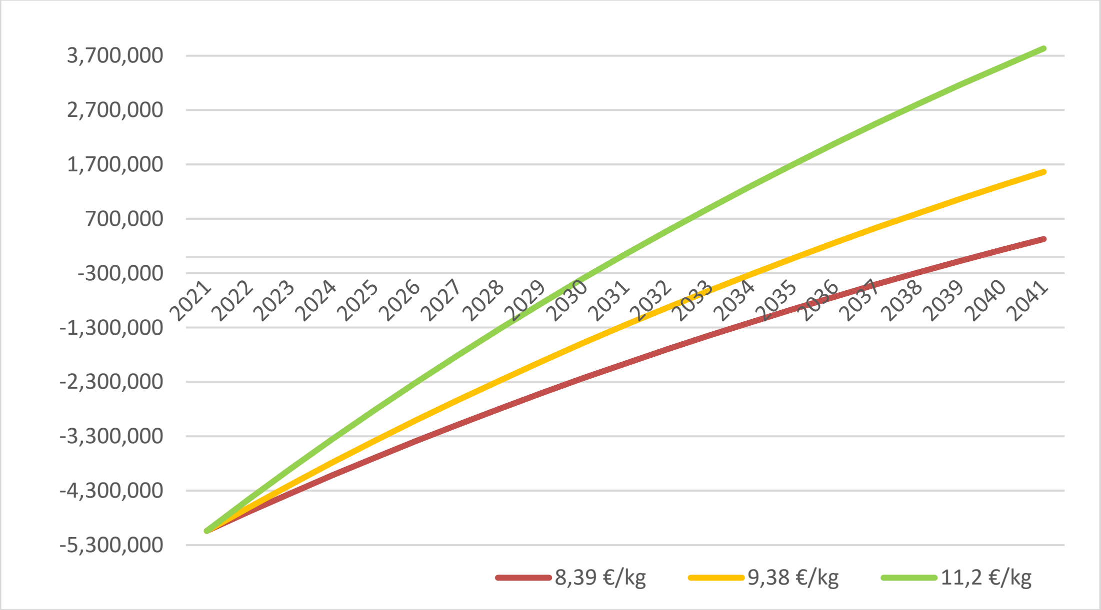


Figure 52 - Configuration 1: cash flows sensitivity analysis for three different hydrogen selling prices

The payback time is the amount of time necessary to gain back the invested capital: inverse proportionality is defined between the selling price and payback time.

$$Selling\ price \propto \frac{1}{PayBack\ Time}$$

Equation 25

Payback time for a selling price of 8.39 €/kg is 19 years, for 9.38 €/kg is 15 years and for 11.2 €/kg is 10 years.

Similarly, LCOE and LCOH for biomethane plant are obtained and various selling price scenarios are compared in the cash flow analysis. In this case, for a selling price of 9.38 €/kg the payback time is 18 years, for 10.3 €/kg it is 14 years and for 12.8 €/kg it is 9 years.

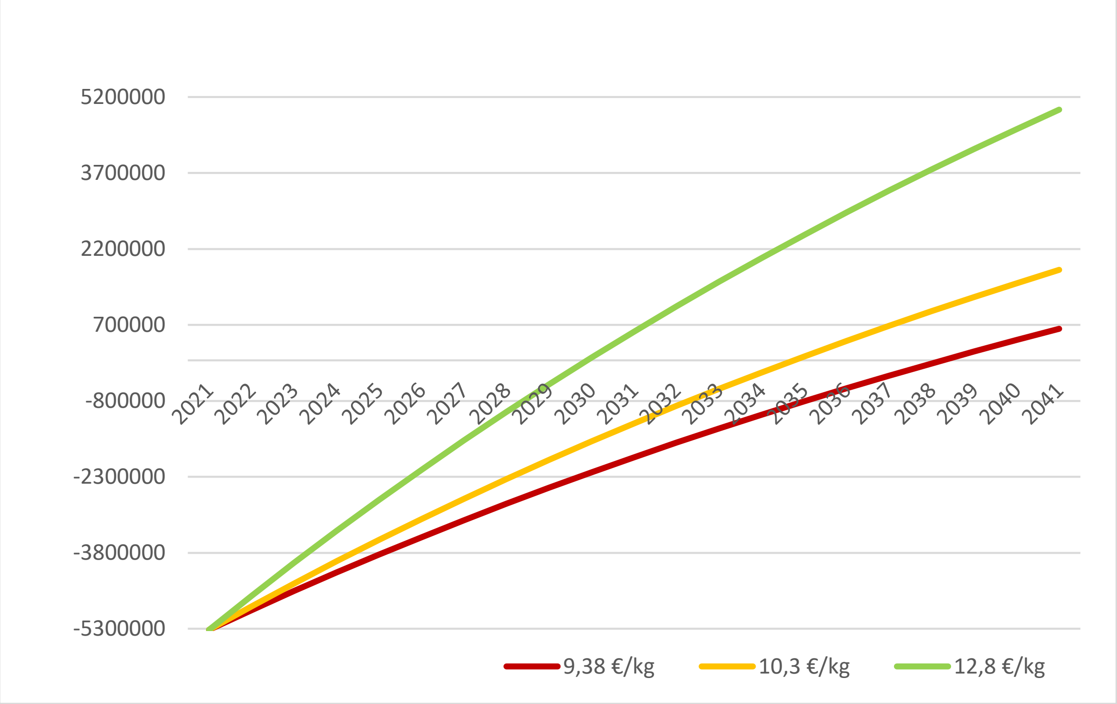


Figure 53 - Configuration 2: cash flow sensitivity analysis for three different hydrogen selling prices

Economically it has been evaluated as feasible with LCOH of 7.6 €/kg (biogas) and 8.3 €/kg (biomethane), LCOE respectively is 228 €/kWh and 250 €/kWh. The plant overall cost is 5.1 million € (biogas) and 5.3 million € (biomethane). Depending on the H_2 selling price the payback time can reach 9 or 10 years, respectively. These values are obtained for hydrogen selling price at the hydrogen refueling station of 11.2 €/kg (biogas) and 12.8 €/kg (biomethane). Lower selling prices are explored too, all the values are consistent with actual prices at HRS. Since the project is built on an island many equipment pieces will arrive with maritime transport: such costs have not been considered in the current analysis.

Both in configuration 1 and 2, NPVs obtained are positive:

	Selling price (€/kg)	NPV	IRR	PBT (years)
Configuration 1	8.39	329,624 €	0.68%	19
	9.38	1,564,916 €	3.04%	15
	11.20	3,505,610 €	6.65%	10
Configuration 2	9.38	623,048 €	1.19%	18
	10.30	1,787,282 €	3.27%	14
	12.80	4,950,962 €	8.26%	9

Table 53 - Summary of NPV and IRR evaluations depending on selling price

Table 50 summarizes NPV and IRR evaluation for both cases: even though all the net present values are positive, the lower selling prices give too low IRR. In red, underlined the unfeasible selling prices and in orange once with higher risk (because IRR is almost equal to discount rate).

In table 54, LCOH and LCOE are reported for first and second configuration:

	LCOH (€/kg)	LCOE (€/kWh)
Configuration 1	7.614	228.42
Configuration 2	8.328	249.84

Table 54 - LCOH and LCOE for both configurations

An important observation must be pointed out on the selling price. The techno-economic analysis evaluated the plant from the anaerobic digester to the hydrogen refueling station, including distribution via trailers. Since has been performed for the construction and operation of the whole plant, the selling price defined is the value that sees the direct consumer at the HRS. According to California Energy Commission report written in 2019, the average hydrogen selling price in dispensing stations is around 16.50 $/kg (14.03 €/kg with 0.85 USD-EUR conversion used in all the document) [70]. Lower selling prices are analyzed in order to verify the economic feasibility even in case the hydrogen price may reduce in the next years: the investment may still result interesting.

However, nowadays hydrogen from renewables is getting cheaper and cheaper, slowly electrolyzer and renewable energy source are scaling-up the production and reducing specific costs: for example, in Taibi's et al. outlook, LCOH from Danish production in 2017 is reported to be around 6 $/kg (5.1 €/kg) for 40% load factor.

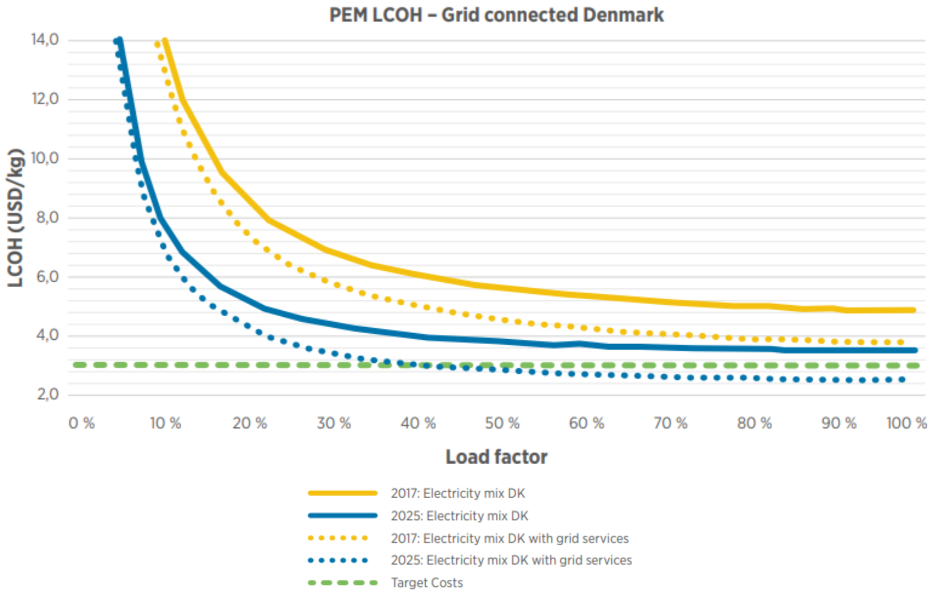


Figure 54 - Levelized cost of hydrogen (LCOH) produced via PEM in 2017 and as expected in 2025 [71]

In general, such plant may be an interesting investment after green hydrogen production plants will be deployed in the area.

Since the plant is very small, lower specific costs may be obtained in case the plant is scaled up: by pairing a hydrogen production facility to a bigger farm, higher production can be obtained. The plant has a high potential since both the anaerobic digestion and the reforming technique are very well developed and diffused: as shows the CAPEX graph, around one third of the investment is linked to the HRS (a novel technology that still has to be scaled up).

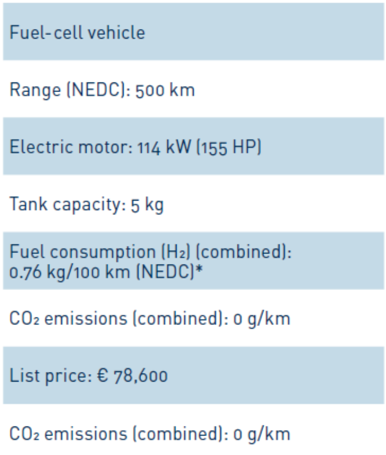
With well-known technologies the OPEX are reduced too and lower cost for maintenance and service are easier to find than for novel technologies. The plant operates thanks to manure produced by cattle in the farm: the source is renewable and the amount of production is stable and continuous (a very interesting feature with respect to others renewable sources).

5.3 End-users vehicle fleet analysis

In order to better define the project and its effects in transportation, a hydrogen vehicle fleet is considered to be completely powered by the plant. Since the daily hydrogen production is almost the same, only the first configuration plant is chosen for these calculations.

Nowadays, the first commercial cars, buses and trucks are under development; due to the low development of the hydrogen network for transportation, only a small number of models are available. Among the most important models, H_2.live introduces Toyata Mirai, Hyundai Nexo, Mercedes-Benz GLF F-Cell, Hyundai IX35, Honda Clarity Fuel Cell and Renault Kangoo ZE H_2 as the most interesting actual solutions. All of them have tank capacity that gravitates around 5 kg (from 4.4 kg to 6.33 kg). Car tank is a storage at 700 bar that aliments PEM fuel cell of around 100-130 kW (depending on the model). For Toyota Mirai, the H_2.live presentation reports hydrogen consumption of 0.76 kg/100 km, but 1 kg/100 km given in the “Study on early business cases for H_2” funded by FCH JU is considered for calculations [55], [72].

The autonomy range is given as 500 km.



Fuel-cell vehicle

Range (NEDC): 500 km

Electric motor: 114 kW (155 HP)

Tank capacity: 5 kg

Fuel consumption (H_2) (combined): 0.76 kg/100 km (NEDC)*

CO_2 emissions (combined): 0 g/km

List price: € 78,600

CO_2 emissions (combined): 0 g/km

Figure 55 - Toyota Mirai technical sheet

From Google Maps and Canary Islands information website bus trips have been found, in La Aldea de San Nicolas two bus lines are present: line 38 goes to Mogan with a 27 km route (two buses will be needed due to the long trip) and line 101 goes to Puerto de San Nicolas at 5.3 km of La Aldea (one bus only is needed for short trip). Hydrogen bus consumption came from the study funded by FCH JU and accounts for 10 kg/100 km [55]. The tables below report the information on average daily mileage and hydrogen consumption according to the bus schedule:

La Aldea - Mogan	Line 38	
Number of buses	2	
Single trip	27	km
Round Trip	54	km
Daily trips	5	
Daily mileage	540	km
Daily H_2 consumption	54	kg/day

Table 55 - Bus line 38 informations

La Aldea - Puerto de San Nicolas	Line 101	
Single trip	5.3	km
Round Trip	10.6	km
Daily trips	4	
Daily mileage	42.4	km
Daily H_2 consumption	4.24	kg/day

Table 56 - Bus line 101 informations

A forklift fleet is considered to be present in the surrounding factories, farms or ports: forklifts hydrogen consumption is reported as 1 kg/day [55] and 30 forklifts are assumed to be present in the area.

Additionally, a catamaran hydrogen powered could be present in future and may need fuel supply: the daily hydrogen consumption has been supposed as 30 kg/day.

With the given information, two scenarios are presented for the daily car consumption evaluation. In the first one, Dalla Chiara's paper is used to obtain the average daily distance is covered by car owners: in the study 34.8 km is given by Politecnico di Torino [73]. Since the study is not taken out specifically for an island a second scenario is developed, to cover lack of capillary public transport (or reduced service) most of the people that live on islands use their own car to move: the average daily distance is doubled.

5.3.1 First scenario

A single car average daily distance is 34.8 km, weekly distance is 243.6 km. FCEVs have 500 kg tanks, it means that H_2 refill must be done at least every two weeks (twice a month). Moreover, last year data from ISTAC report 7,508 persons populating La Aldea de San Nicolas: 100 cars are assumed to circulate in the area in the future [74]. According to these assumptions, a single car consumes around 0.35 kg/day and 2.4 kg/week, the total fleet would need around 35 kg/day and around 245 kg/week.

Cars	Single car		100 cars	
Average daily mileage	34.8	km	3,480	km
Daily H_2 consumption	0.348	kg/day	34.8	kg/day
Weekly H_2 consumption	2.436	kg/week	243.6	kg/week

Table 57 - Scenario 1: average daily and weekly cars H_2 consumption

The total daily consumption for the first scenario would be resume in table 58 and figure 56:

SCENARIO 1

Type	Consumption	
Cars	34.8	kg/day
Buses	58.2	
Forklifts	30	
Ship	30	
Total	153.0	

Table 58 – Scenario 1: complete fleet H_2 consumption

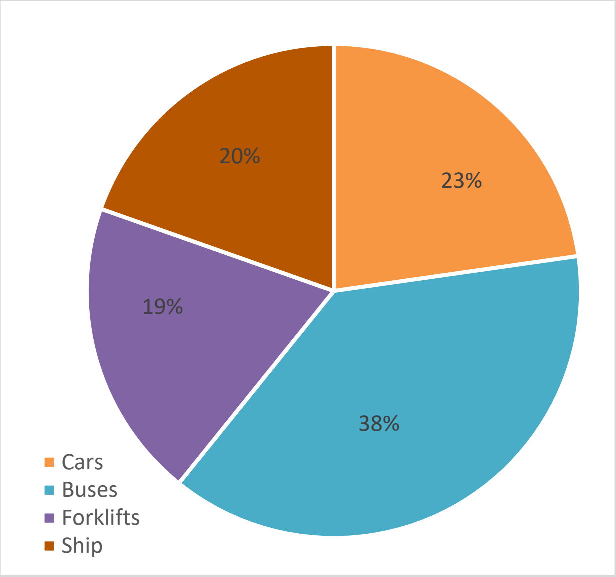


Figure 56 - Scenario 1: daily consumption distribution

5.3.2 Second scenario

Similarly, average consumption is doubled: every single car is assumed to cover almost 70 km/day with 0.69 kg/day and 4.87 kg/week.

Cars	Single car		100 cars	
Average daily mileage	69.6	km	6,960	km
Daily H_2 consumption	0.696	kg/day	69.6	kg/day
Weekly H_2 consumption	4.87	kg/week	48.2	kg/week

Table 59 - Scenario 2: average daily and weekly cars H_2 consumption

The results are then reported in the table and figure below:

SCENARIO 2

Type	Consumption	
Cars	69.6	kg/day
Buses	58.2	
Forklifts	30	
Ship	30	
Total	187.8	

Table 60 - Scenario 2: complete fleet H_2 consumption

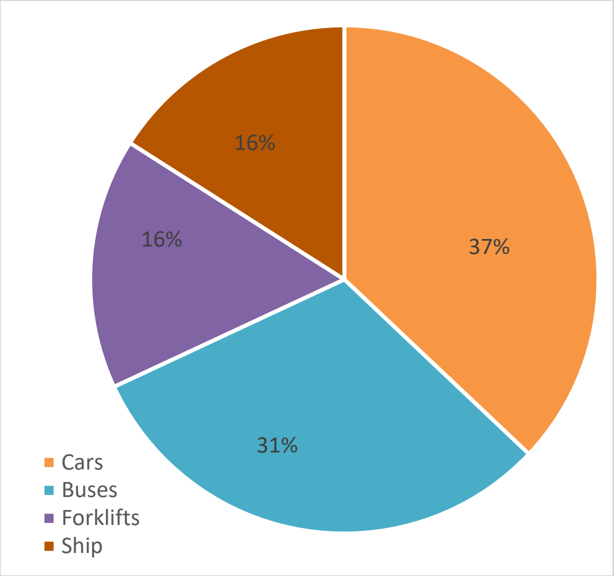


Figure 57 - Scenario 2: daily consumption distribution

In compliance with the previous evaluations, the hydrogen produced by the plant should be high enough to grant the daily consumption by the whole fleet estimated for La Aldea. With nominal hydrogen production, the daily mass flowrate overcomes the request for both scenarios and still allows to refill storages that may have been emptied in the days before.

The plant operated in nominal conditions would be able to supply a vehicle fleet of 100 cars, 3 buses, 30 forklifts and one catamaran. Two scenarios have been evaluated depending on cars daily average covered distance: in first case, 35 km are covered and, second case, almost 70 km. For first scenario, 100 cars are powered by 34.8 kgH_2/day and in second case 69.6 kgH_2/day are necessary. The buses, forklifts and ship fleet has around 120 kgH_2/day request.

Additionally, the number of cows necessary to power a single car or 100 cars is calculated for both scenarios:

Single car	Scenario 1	Scenario 2	
Daily consumption	0.348	0.696	kg
Daily hydrogen production	229.78		kg
Daily car percentage	0.15%	0.30%	
Daily cattle-to-car	1.5 (at least 2)	3	cows
100 cars	Scenario 1	Scenario 2	
Daily consumption	34.8	69.6	kg
Daily hydrogen production	229.78		kg
Daily car percentage	15.14%	30.29%	
Daily cattle-to-car	151	303	cows

Table 61 - Single car and 100 cars minimum cattle calculations

The minimum number of cows necessary to power a car is 1.5 (so, at least 2 cows) for first scenario and 3 for second scenario. The same result is obtained for the 100 cars to be powered: for first scenario, 151 cattle are needed and 303 for second scenario.

6 CONCLUSIONS

6.1 Conclusions from the work

The techno-economic feasibility of two possible configurations of a manure-to-hydrogen conversion plant have been explored and the following conclusions have been drawn:

1. Both configurations designed are evaluated as feasible from the technical point of view: hydrogen for mobility purposes has been obtained. From economic analysis, positive results have been obtained because hydrogen from a renewable source for mobility purposes usually has higher remunerations.
2. However, nowadays green hydrogen from electrolysis is leading to better LCOH. Once hydrogen production from renewables is established in this location, the project can be considered as an available alternative thanks to high quantity manure daily production coming from the farm.
3. Configuration 1 is the most interesting from the economic point of view, this is due to lower refining of the middle-product during the processes that leads to cheaper solutions. Economic indicators have been used to compare biogas-based solution, with biomethane one:
    a. Payback time is lower because of lower investment.
    b. LCOH obtained is smaller: it allows to sell hydrogen at lower prices and be more competitive on the market.
    c. CAPEX is lower: configuration 1 is financed with 5.1 million € and configuration 2 needs 5.3 million €.
    d. In both configurations, IRR is not high enough for low selling prices (8.38 €/kg for configuration 1 and 9.38 €/kg for configuration 2): by increasing selling prices, the investment become profitable.
    e. NPV is positive in all cases explored, but higher values are obtained with configuration 1.

4. In general, a considerably high hydrogen demand for mobility purposes is required to justify such investment. In an optimistic future in which hydrogen mobility will be ordinary, such project may become part of the hydrogen supply chain for hydrogen refueling stations.

5. Thus, the recommendation would be a project by phases, starting with a small pilot plant to refuel a few vehicles in a project with public funding to then upscale to use all the manure. With such approach, most of hydrogen valleys are built today to avoid risks of high CAPEX investments.

6.2 Recommendations for future works

The following aspects has been detected as potential research lines:

1. Manure chemical composition could be studied in order to obtain real data and have higher certainties on biogas yield.
2. Further evaluations can be done to implement carbon capture, utilization and storage (CCUS) to reduce emissions and, since the plant has quite pure CO_2, create a new by-product that may be sold.
3. Highly detailed optimization of the plant circuits could be performed in order to reduce consumption and use high-temperature heat used in hydrogen production section.

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8 APPENDIX

8.1 Cash flow evaluation tables

Case 1	Hydrogen price	8.39	€/kg

Year	Cost	Revenues	Cash Flow	Time	Discount Rate	Present Cash Flow	Cumulated Cash Flow
2021	-5,037,988		-5,037,988	0	1.000	-5,037,988	-5,037,988
2022	-342,879	703,667	360,788	1	0.971	350,279	-4,687,708
2023	-342,879	703,667	360,788	2	0.943	340,077	-4,347,631
2024	-342,879	703,667	360,788	3	0.915	330,172	-4,017,459
2025	-342,879	703,667	360,788	4	0.888	320,555	-3,696,904
2026	-342,879	703,667	360,788	5	0.863	311,219	-3,385,685
2027	-342,879	703,667	360,788	6	0.837	302,154	-3,083,531
2028	-342,879	703,667	360,788	7	0.813	293,354	-2,790,178
2029	-342,879	703,667	360,788	8	0.789	284,809	-2,505,368
2030	-342,879	703,667	360,788	9	0.766	276,514	-2,228,855
2031	-342,879	703,667	360,788	10	0.744	268,460	-1,960,395
2032	-342,879	703,667	360,788	11	0.722	260,641	-1,699,754
2033	-342,879	703,667	360,788	12	0.701	253,049	-1,446,704
2034	-342,879	703,667	360,788	13	0.681	245,679	-1,201,025
2035	-342,879	703,667	360,788	14	0.661	238,523	-962,502
2036	-342,879	703,667	360,788	15	0.642	231,576	-730,926
2037	-342,879	703,667	360,788	16	0.623	224,831	-506,095
2038	-342,879	703,667	360,788	17	0.605	218,283	-287,813
2039	-342,879	703,667	360,788	18	0.587	211,925	-75,888
2040	-342,879	703,667	360,788	19	0.570	205,752	129,865
2041	-342,879	703,667	360,788	20	0.554	199,759	329,624

NPV	IRR
329,624	0.68%

Case 1	Hydrogen price	9.38	€/kg

Year	Cost	Revenues	Cash Flow	Time	Discount Rate	Present Cash Flow	Cumulated Cash Flow
2021	-5,037,988		-5,037,988	0	1.000	-5,037,988	-5,037,988
2022	-342,879	786,698	443,819	1	0.971	430,892	-4,607,096
2023	-342,879	786,698	443,819	2	0.943	418,342	-4,188,754
2024	-342,879	786,698	443,819	3	0.915	406,157	-3,782,597
2025	-342,879	786,698	443,819	4	0.888	394,327	-3,388,270
2026	-342,879	786,698	443,819	5	0.863	382,842	-3,005,428
2027	-342,879	786,698	443,819	6	0.837	371,691	-2,633,736
2028	-342,879	786,698	443,819	7	0.813	360,865	-2,272,871
2029	-342,879	786,698	443,819	8	0.789	350,355	-1,922,516
2030	-342,879	786,698	443,819	9	0.766	340,150	-1,582,366
2031	-342,879	786,698	443,819	10	0.744	330,243	-1,252,123
2032	-342,879	786,698	443,819	11	0.722	320,624	-931,499
2033	-342,879	786,698	443,819	12	0.701	311,286	-620,213
2034	-342,879	786,698	443,819	13	0.681	302,219	-317,994
2035	-342,879	786,698	443,819	14	0.661	293,417	-24,578
2036	-342,879	786,698	443,819	15	0.642	284,870	260,293
2037	-342,879	786,698	443,819	16	0.623	276,573	536,866
2038	-342,879	786,698	443,819	17	0.605	268,518	805,383
2039	-342,879	786,698	443,819	18	0.587	260,697	1,066,080
2040	-342,879	786,698	443,819	19	0.570	253,104	1,319,184
2041	-342,879	786,698	443,819	20	0.554	245,732	1,564,916

NPV	IRR
1,564,916	3.04%

Case 1	Hydrogen price	11.2	€/kg

Year	Cost	Revenues	Cash Flow	Time	Discount Rate	Present Cash Flow	Cumulated Cash Flow
2021	-5,037,988		-5,037,988	0	1.000	-5,037,988	-5,037,988
2022	-342,879	939,341	596,462	1	0.971	579,089	-4,458,899
2023	-342,879	939,341	596,462	2	0.943	562,222	-3,896,677
2024	-342,879	939,341	596,462	3	0.915	545,847	-3,350,830
2025	-342,879	939,341	596,462	4	0.888	529,948	-2,820,881
2026	-342,879	939,341	596,462	5	0.863	514,513	-2,306,368
2027	-342,879	939,341	596,462	6	0.837	499,527	-1,806,841
2028	-342,879	939,341	596,462	7	0.813	484,978	-1,321,863
2029	-342,879	939,341	596,462	8	0.789	470,852	-851,010
2030	-342,879	939,341	596,462	9	0.766	457,138	-393,872
2031	-342,879	939,341	596,462	10	0.744	443,824	49,951
2032	-342,879	939,341	596,462	11	0.722	430,897	480,848
2033	-342,879	939,341	596,462	12	0.701	418,346	899,194
2034	-342,879	939,341	596,462	13	0.681	406,161	1,305,356
2035	-342,879	939,341	596,462	14	0.661	394,331	1,699,687
2036	-342,879	939,341	596,462	15	0.642	382,846	2,082,533
2037	-342,879	939,341	596,462	16	0.623	371,695	2,454,228
2038	-342,879	939,341	596,462	17	0.605	360,869	2,815,097
2039	-342,879	939,341	596,462	18	0.587	350,358	3,165,456
2040	-342,879	939,341	596,462	19	0.570	340,154	3,505,610
2041	-342,879	939,341	596,462	20	0.554	330,246	3,835,856

NPV	IRR
3,505,610	6.65%

Case 2	Hydrogen price	9.38	€/kg

Year	Cost	Revenues	Cash Flow	Time	Discount Rate	Present Cash Flow	Cumulated Cash Flow
2021	-5321132.9		-5321132.9	0	1.000	-5321132.9	-5321132.9
2022	-398316.7	797859.0	399542.3	1	0.971	387905.2	-4933227.7
2023	-398316.7	797859.0	399542.3	2	0.943	376606.9	-4556620.8
2024	-398316.7	797859.0	399542.3	3	0.915	365637.8	-4190983.0
2025	-398316.7	797859.0	399542.3	4	0.888	354988.2	-3835994.8
2026	-398316.7	797859.0	399542.3	5	0.863	344648.7	-3491346.1
2027	-398316.7	797859.0	399542.3	6	0.837	334610.4	-3156735.7
2028	-398316.7	797859.0	399542.3	7	0.813	324864.5	-2831871.2
2029	-398316.7	797859.0	399542.3	8	0.789	315402.4	-2516468.9
2030	-398316.7	797859.0	399542.3	9	0.766	306215.9	-2210252.9
2031	-398316.7	797859.0	399542.3	10	0.744	297297.0	-1912955.9
2032	-398316.7	797859.0	399542.3	11	0.722	288637.9	-1624318.1
2033	-398316.7	797859.0	399542.3	12	0.701	280230.9	-1344087.1
2034	-398316.7	797859.0	399542.3	13	0.681	272068.9	-1072018.3
2035	-398316.7	797859.0	399542.3	14	0.661	264144.5	-807873.7
2036	-398316.7	797859.0	399542.3	15	0.642	256451.0	-551422.7
2037	-398316.7	797859.0	399542.3	16	0.623	248981.6	-302441.2
2038	-398316.7	797859.0	399542.3	17	0.605	241729.7	-60711.5
2039	-398316.7	797859.0	399542.3	18	0.587	234689.0	173977.5
2040	-398316.7	797859.0	399542.3	19	0.570	227853.4	401830.9
2041	-398316.7	797859.0	399542.3	20	0.554	221216.9	623047.8

NPV	IRR
623,048	1.19%

Case 2	Hydrogen price	10.3	€/kg

Year	Cost	Revenues	Cash Flow	Time	Discount Rate	Present Cash Flow	Cumulated Cash Flow
2021	-5321132.9		-5321132.9	0	1.000	-5321132.9	-5321132.9
2022	-398316.7	876113.9	477797.1	1	0.971	463880.7	-4857252.2
2023	-398316.7	876113.9	477797.1	2	0.943	450369.6	-4406882.5
2024	-398316.7	876113.9	477797.1	3	0.915	437252.1	-3969630.5
2025	-398316.7	876113.9	477797.1	4	0.888	424516.6	-3545113.9
2026	-398316.7	876113.9	477797.1	5	0.863	412152.0	-3132961.9
2027	-398316.7	876113.9	477797.1	6	0.837	400147.6	-2732814.3
2028	-398316.7	876113.9	477797.1	7	0.813	388492.8	-2344321.5
2029	-398316.7	876113.9	477797.1	8	0.789	377177.5	-1967144.0
2030	-398316.7	876113.9	477797.1	9	0.766	366191.7	-1600952.3
2031	-398316.7	876113.9	477797.1	10	0.744	355525.9	-1245426.4
2032	-398316.7	876113.9	477797.1	11	0.722	345170.8	-900255.5
2033	-398316.7	876113.9	477797.1	12	0.701	335117.3	-565138.2
2034	-398316.7	876113.9	477797.1	13	0.681	325356.6	-239781.6
2035	-398316.7	876113.9	477797.1	14	0.661	315880.2	76098.6
2036	-398316.7	876113.9	477797.1	15	0.642	306679.8	382778.4
2037	-398316.7	876113.9	477797.1	16	0.623	297747.4	680525.7
2038	-398316.7	876113.9	477797.1	17	0.605	289075.1	969600.9
2039	-398316.7	876113.9	477797.1	18	0.587	280655.5	1250256.3
2040	-398316.7	876113.9	477797.1	19	0.570	272481.0	1522737.4
2041	-398316.7	876113.9	477797.1	20	0.554	264544.7	1787282.1

NPV	IRR
1,787,282	3.268%

Case 2	Hydrogen price	12.8	€/kg

Year	Cost	Revenues	Cash Flow	Time	Discount Rate	Present Cash Flow	Cumulated Cash Flow
2021	-5321132.9		-5321132.9	0	1.000	-5321132.9	-5321132.9
2022	-398316.7	1088762.9	690446.1	1	0.971	670336.1	-4650796.8
2023	-398316.7	1088762.9	690446.1	2	0.943	650811.7	-3999985.1
2024	-398316.7	1088762.9	690446.1	3	0.915	631856.0	-3368129.1
2025	-398316.7	1088762.9	690446.1	4	0.888	613452.5	-2754676.6
2026	-398316.7	1088762.9	690446.1	5	0.863	595584.9	-2159091.7
2027	-398316.7	1088762.9	690446.1	6	0.837	578237.8	-1580854.0
2028	-398316.7	1088762.9	690446.1	7	0.813	561395.9	-1019458.1
2029	-398316.7	1088762.9	690446.1	8	0.789	545044.6	-474413.5
2030	-398316.7	1088762.9	690446.1	9	0.766	529169.5	54756.0
2031	-398316.7	1088762.9	690446.1	10	0.744	513756.8	568512.7
2032	-398316.7	1088762.9	690446.1	11	0.722	498793.0	1067305.7
2033	-398316.7	1088762.9	690446.1	12	0.701	484265.0	1551570.8
2034	-398316.7	1088762.9	690446.1	13	0.681	470160.2	2021731.0
2035	-398316.7	1088762.9	690446.1	14	0.661	456466.2	2478197.2
2036	-398316.7	1088762.9	690446.1	15	0.642	443171.1	2921368.3
2037	-398316.7	1088762.9	690446.1	16	0.623	430263.2	3351631.5
2038	-398316.7	1088762.9	690446.1	17	0.605	417731.3	3769362.8
2039	-398316.7	1088762.9	690446.1	18	0.587	405564.3	4174927.1
2040	-398316.7	1088762.9	690446.1	19	0.570	393751.8	4568678.9
2041	-398316.7	1088762.9	690446.1	20	0.554	382283.3	4950962.2

NPV	IRR
4,950,962	8.258%