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STORE&GO

Innovative large-scale energy storage technologies and Power-to-Gas concepts after optimisation

D8.3

Report on the costs involved with PtG technologies and their potentials across the EU

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1.7	2018-04-12	C van Leeuwen	Final version including last remarks of Frank Graf, Simon Verleger and Jachin Gorre

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Executive Summary

This report presents an overview of *current* costs of power-to-gas (PtG) plants that produce methane, including investment costs, operational costs and feedstock costs. The costs are estimated based on literature sources and data from the STORE&GO demonstration plants that are currently built in Falkenhagen (Germany), Solothurn (Switzerland) and Troia (Italy). The report first gives an overview of the investment and operational costs for the different PtG plant components: the electrolyser, hydrogen storage, methanation reactor, CO_2 storage, compressors, pipelines, gas grid injection, re-fuelling station and oxygen and heat capture. Next, total costs are determined for regular-sized PtG plants according to nowadays standards (1 – 5 MW electrical input). Feedstock costs – water, electricity and CO_2 – are also discussed.

The costs presented in this report will serve as a starting point for further analysis of future cost developments of the PtG technology due to learning effects and economies of scale in STORE&GO deliverable 7.5[a], and will also be used as an input for PtG business calculations in STORE&GO deliverable 8.6[b]. Although this report presents already some costs of the three STORE&GO demonstration plants, not all costs could be provided at this point in the project since the plants are still being constructed. A detailed evaluation of the costs of the three demonstration plants will be given later in STORE&GO deliverable 5.9[c].

With all costs combined (investment, operational and feedstock costs), the methane production costs in a PtG plant can be calculated and compared with natural gas prices to assess the potential of the technology. The methane production costs of PtG plants were calculated to be 1.95 €/kg for full time production of the plant and these costs can be roughly divided into 43% electricity costs, 37% CAPEX, 13% OPEX and 7% costs for CO_2. Operating the plant only during hours with low electricity prices was found not to improve the overall business case of the plant due to the higher burden of the CAPEX and OPEX on the produced methane. Costs for water were found to be negligible in all cases (adding a maximum of 0.3% to the total costs).

Comparing the production costs of methane from PtG plants to the current costs of natural gas (roughly 0.30 €/kg) it is clear that PtG cannot compete with fossil gas today. We illustrate the effect of a decrease by half of the electrolyser investment costs – the most expensive component of the PtG plant in terms of investment costs – and find that production costs of methane decrease from 1.95 €/kg to 1.63 €/kg. Although this is a significant cost reduction (16%) it is – by far – not sufficient to bridge the gap to the price of natural gas.

Many things can change in the (near) future that would improve the business case of PtG. A further decrease in the electrolyser investment costs is possible, just as a decrease in costs of other plant components. The electricity prices could also change: a reduction in the average price but also a different pattern with more hours with low prices would be beneficial for PtG. It is important to remark, however, that the electricity prices used in the calculations in this report are already the lowest prices that have been observed in recent years (Germany 2016) and that further reductions might be feasible in the short-run but not in the long-run.

The revenue of the produced gas ultimately defines whether the PtG plant can operate economically or not. The methane has to compete with natural gas, but the revenues could be higher due to the

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[a] D7.5: Report on experience curves and economies of scale. Due date: 31 October 2018

[b] D8.6: Report on the optimal time profile and operation of the conversion technology during a representative year, in the perspective of available storage capacities. Due date: 31 December 2018

[c] D5.9: Final report on evaluation of technologies and processes. Due date: 29 February 2020.

green character of the gas, as opposed to the fossil nature of natural gas. The market perspective of green gases is a topic that will be further investigated in deliverable 8.5[d].

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[d] D8.5: Study describing the short, medium and long-term perspectives of various market segments for “green gases”. Due date: 31 October 2018

1 Introduction

In a power-to-gas (PtG) plant, electricity is converted into hydrogen using an electrolyser. It can further be converted into methane using a methanation reactor. Producing hydrogen or methane in a PtG plant involves many different costs, including investment costs, operational costs and feedstock costs. This report provides an overview of all components that can be of importance for a PtG plant and provides estimates of the *current* costs of these components.

Since the PtG technology is relatively new and fully in development, many costs (such as those of the electrolyser) are expected to decrease in the (near) future due to learning effects and economies of scale. These future effects will not be discussed in this report but will be the main topic of STORE&GO deliverable 7.5[e]. This report will thereby serve as a starting point for investigating these future cost reduction potentials.

Figure 1 gives an overview of the basic cost components associated with a PtG plant. The power-to-hydrogen plant is shown within the black dotted box and consists of an electrolyser stack accompanied with equipment such as a power conversion system, water treatment and gas purification system. Together with other things such as piping, structure housing and measurement equipment, these form the balance of plant (BoP) of the electrolyser, and everything together is referred to as the electrolyser system. The hydrogen is stored in a hydrogen storage facility for which a compressor is usually needed. In case the produced hydrogen needs to be converted into methane, additional equipment is needed in the form of a methanation reactor with its BoP and a CO_2 storage tank with a compressor. The different feedstock needed in PtG plants (electricity, water and CO_2) are indicated in orange in the figure, whereas the products (hydrogen, methane, oxygen and heat) are marked in green.

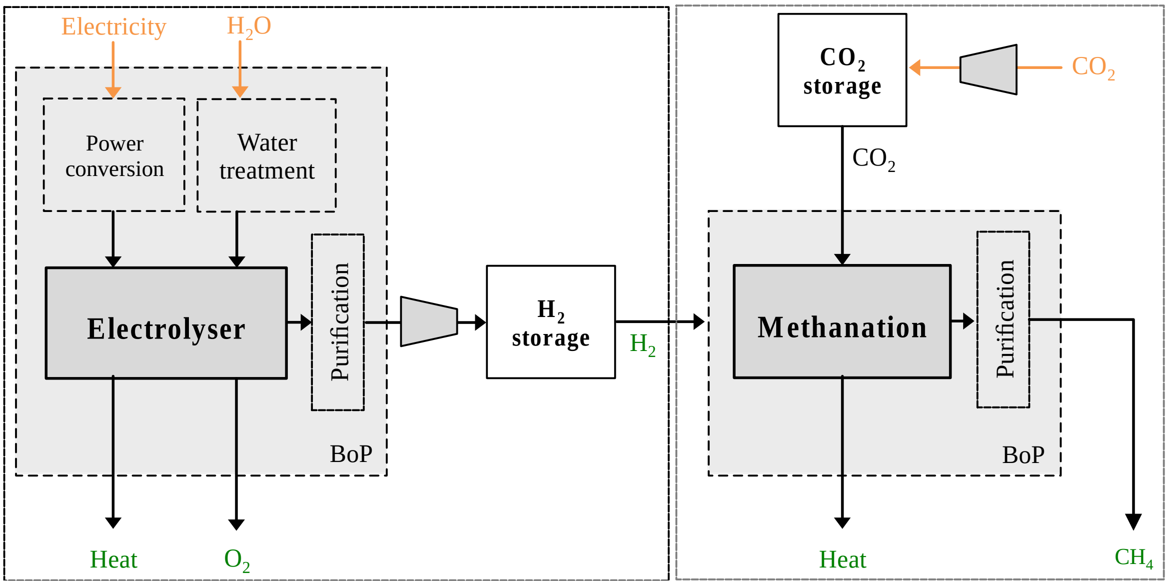

Figure 1: Overview of a PtG plant producing hydrogen or methane. Main investment costs for the different plant components are numbered in orange. The feedstock electricity, water and CO_2 are marked in orange.

Depending on the final application and destination of the produced gas, additional investments could be needed, that are not shown in Figure 1. Examples are pipelines, a gas grid injection station or refuelling station. Besides the investment costs for all basic components of the plant, additional costs will be made for planning, design, preparation and installation of the plant. These costs will increase

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[e] D7.5: Report on experience curves and economies of scale. Due date: 31 October 2018

the total investment costs. Once the plant is in operation, operational costs will be made for personnel and maintenance.

The different cost estimates presented in this report are made based on available literature sources as well as on data from the three STORE&GO demonstration plants that are built at the moment in Falkenhagen (Germany), Solothurn (Switzerland) and Troia (Italy). At this point in the project, however, not all costs of the demonstration sites are known. A more detailed evaluation of investment and operational costs of the demonstration plants will be provided in the end of the STORE&GO project, in Deliverable 5.9[f].

The structure of this report is as follows. Chapter 2 will provide an overview of the investment and yearly fixed operational costs of PtG plants, including the costs for all basic components as presented in Figure 1, as well as an estimation of the total costs for a PtG plant that is producing methane. As was said before, all costs presented refer to the current situation. In chapter 3, feedstock costs are discussed, including the costs for water, electricity and CO_2. Chapter 4 discusses the presented costs and calculates the methane production costs in PtG plants to assess the potential of the technology.

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[f] D5.9: Final report on evaluation of technologies and processes. Due date: 29 February 2020.

2 Investment and fixed operational costs of PtG plants

This chapter discusses the investment and fixed operational costs of PtG plants by first estimating these costs for the basic components of PtG plants (section 2.1) and afterwards determining the total costs of a PtG plant that produces methane (section 2.2). The costs presented here are estimates of the *current* situation and are based on literature sources as well as on data provided by the three STORE&GO demonstration sites. Economies of scale are not discussed here – this will be done extensively in STORE&GO deliverable 7.5[g] - but it is clear that for most components the size does influence the costs. The costs presented here refer to an averaged-sized demonstration plant (for nowadays standards) of about 1 – 5 MW electrical input.

2.1 Investment and operational costs basic components PtG plants

This section will describe the investment and operational costs of all basic components of a PtG plant. It follows Figure 1 and includes the electrolyser stack and accompanying BoP (together the electrolyser system), the hydrogen storage, the methanation reactor and accompanying BoP and the CO_2 storage. Additionally, investment costs for compressors, pipelines, a gas grid injection station, a refuelling station and oxygen and heat capture facilities are discussed.

2.1.1 Electrolyser system

The heart of every PtG plant is the electrolyser. In the electrolyser, water is split into the components oxygen (O_2) and hydrogen (H_2) by using electricity according to equation 1:

$$2H_2O \rightarrow 2H_2 + O_2 \qquad \textbf{Equation 1}$$

Three techniques are currently available for water electrolysis: alkaline electrolysis (AEC), proton exchange membrane electrolysis (PEMEC) and solid oxide electrolysis cells (SOEC). Alkaline and PEM electrolysers operate at relatively low temperature. Alkaline electrolysis was originally the most highly developed and cheapest technology and therefore most commonly used (Gahleitner, 2013; Holladay et al., 2009). Nowadays, however, PEM electrolysers are a serious alternative for alkaline electrolysers and they are used in several PtG pilot plants (Gahleitner, 2013; Kopp et al., 2017). PEM electrolysers can reach higher efficiencies and can deal with fast load changes, which can be very beneficial in PtG applications (Gahleitner, 2013). In an expert elicitation study published by (Schmidt et al., 2017) it was found that a majority of experts expected PEM electrolysis to become the most important electrolysis technique by 2030 due to its superior characteristics for intermittent operation. Before this would lead to a commercial advantage over alkaline electrolysis, however, more experience is required and alkaline electrolysis was expected to remain the most important technology in the coming years.

SOEC is different from the other two technologies as it operates at high temperature and uses steam instead of water. The technology is currently the least developed of the three and not yet widely applied on a commercial scale. The most important advantage of SOEC is the low electricity demand and thus potential high efficiency (Bičáková and Straka, 2012; Götz et al., 2016; Ursua et al., 2012; Zahid et al., 2010). The main problems currently associated with the technology are fast material degradation, limited long-term stability due to the high operating temperatures and high capital costs (Götz et al., 2016; Schmidt et al., 2017; Ursua et al., 2012). Current research is focussing on reducing the material degradation and some promising results can be found already (e.g. (Schefold et al., 2017)). An interesting option is the coupling of SOEC with methanation where the heat from the methanation process is used for the vaporization of water before the electrolysis (Giglio et al., 2018; 2015a; 2015b; HELMETH, 2018). Although SOEC could become a very important competitor for

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[g] D7.5: Report on experience curves and economies of scale. Due date: 31 October 2018

alkaline and PEM electrolysers in the future – especially in combination with methanation – the technology is not relevant when assessing current PtG plants. For that reason, only alkaline and PEM electrolysers are considered in this report. Within STORE&GO, two out of the three demonstration plants use an alkaline electrolyser (Germany, Italy) whereas the third uses a PEM electrolyser (Switzerland).

When determining the investment costs of electrolysers, it is important to specify what is included in the costs and whether this is only the electrolyser stack or the complete electrolyser system – including not only the electrolyser stack but also water and gas treatment, power conversion, structure housing, piping and measurement equipment (the BoP). Investment costs are usually expressed in terms of costs per electrical input (€/kW$_{el}$). To assess the business case of PtG plants, also the efficiency of the electrolyser (system) is needed, just as the lifetime of the electrolyser (stack) and the operational costs.

Investment costs

The investment costs of the different types of electrolyser technologies were analysed in an extensive literature screening. The literature review is facing with two problems. On the one hand, currently not much reliable data on costs is available: sources often suffer from a lack of information about the system boundary, year of installation, rated power and source of the reported costs. On the other hand, the available data in the literature have a wide range and is in most cases not up to date. In this report we only consider papers where the costs are based on offers and price inquiries, manufacturers- and expert-elicitations. Literature sources, where only assumptions are made or only other literature sources are summarized, are not taken into account. In a lot of papers costs of electrolysers are given, but they are from papers that refer again to other papers and so on. A lot of calculations in the field of PtG are based on the paper of (Smolinka et al., 2011). However, the data in this paper are not quite up to date (from the years 2002 until 2009). The literature research on the costs of electrolysers has not been completed yet and is being continually updated. Here, we present the first results of the literature review. We only assess current costs and do not predict costs for the future. An update of the literature review will be given in deliverable 7.5[h] that will also include future cost estimates.

In the – earlier mentioned – paper of (Smolinka et al., 2011), specific investment costs for alkaline pressure-less and pressurised electrolysers were calculated on the basis of offers and price requests from the years 2002 to 2009. Depending on the hydrogen production rate (up to 500 m³/h (NTP) which is ~ 2.5 MW) the specific investment costs range from around 750 €/kW$_{el}$ to 6,000 €/kW$_{el}$. The costs for PEMEC were also determined, but hardly any prices were available for electrolysers with a production rate greater than 10 m³/h (equals to app. 50 kW$_{el}$), since no electrolysers existed on this scale. The investment costs in the range of 0.5 to 6 m³/h (NTP) (2.5 – 3 kW$_{el}$) originate from a telephone inquiry with a manufacturer and those in the range of 50 to 200 Nm³/h (250 – 1,000 kW$_{el}$) from theoretical calculations. Depending on the hydrogen production rate of the PEMEC, the specific investment costs are approximately 900 €/kW$_{el}$ (for 200 m³/h (NTP) or ~1 MW) up to 10,000 €/kW$_{el}$ (for 0.4 m³/h (NTP) or ~2 kW). For the calculation of current hydrogen production costs that are carried out in the paper, the authors used specific investment costs of about 2,500 €/kW for a PEMEC (30 m³/h (NTP) or ~150 kW) and 1,000 €/kW for an AEC (500 m³/h (NTP) or ~2.5 MW).

In (Steinmüller et al., 2014) a system analysis of the PtG technology was carried out. One chapter deals with investment costs of the AEC and PEMEC system. It is mentioned, that the indication of current specific investment costs is very difficult, because the system costs primarily depend on the purpose of the field of application. For this reason, the given costs must be seen as guideline values.

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[h] D7.5: Report on experience curves and economies of scale. Due date: 31 October 2018

The specific system costs for AEC and PEMEC are based on price information from electrolyser manufacturers in the year 2014. The specific investment costs for an AEC with a hydrogen production rate of about 100 m^3/h (~0.5 MW) (NTP) can be estimated at about 1,800 €/kW_{el}. The costs drop to 1,200 €/kW_{el} for an electrolyser with 500 m^3/h (NTP) (~2.5 MW). By way of comparison, PEMEC have higher costs for a similar production rate of about 3,500 €/kW_{el} (0.1 MW) and 1,750 €/kW_{el} (1 MW). According to various manufacturers and research institutes, the costs of the entire electrolysis system can be approximately divided into 50 % stack costs (AEC: 40 – 50 %, PEMEC 50-60 %), 10 – 20 % power electronics and 30-40 % remaining costs (BoP).

In (Bertuccioli et al., 2014) an overview of the current status and a forecast for the development of alkaline and PEM electrolysis technology is given. The key figures for the electrolysers are from various literature sources (2010 – 2013) as well as presentations, reports from the US Department of Energy and data sheets from manufacturers. The reported costs are the total costs of an electrolyser system, taking into account the stack, gas management, water management, control and energy supply (transformers, rectifiers). For alkaline electrolyser systems, current costs are reported to fall in the range of 1,000 – 1,200 €/kW for electrolyser with a rated power of 1.8 – 5,300 kW. As a central case, costs of 1,100 €/kW are assumed. PEM electrolyser systems are reported to fall in the range of 1,860 – 2,320 €/kW (rated power of 0.2 – 1,150 kW) with 2,090 €/kW as a central case.

(Energinet dkEnergi Styrelsen, 2012) publish key data on energy generation technologies and thus also on electrolysis technologies at regular intervals. The data are from well-founded and public sources as well as from expert information. The specific investment costs for alkaline electrolysers with a nominal power of less than 3.4 MW are specified with 1,400 €/kW_{el} in the year 2015. Costs for a PEM electrolyser are estimated to be very high at 6,000 €/kW_{el}, however, with a very low rated power of 45 kW.

The data given in (Schenuit et al., 2016) are based on assessments by scientific actors and operators of existing pilot plants. Manufacturers can already offer large alkaline electrolysers for less than 1,000 €/kW in the year 2016. The PEM technology is currently still only produced in smaller sizes and therefore significantly more expensive with costs of around 2,000 €/kW.

(de Bucy, 2016) addresses the economic potential of Power-to-X applications. The costs for the electrolysis technologies are derived from current literature sources. The current specific investment costs for AEC are stated as 2,000 €/kW (rated power 500 kW), 1,500 €/kW (rated power 1 MW) and 1,000 €/kW (rated power 10 MW). The costs include BoP, transport, installation and commissioning. Current investment costs for PEM electrolysers are not given. It is reported that they are higher than those of alkaline electrolysers but that it is expected that costs will decrease fast in the coming years, as PEM electrolyser manufacturers are very active in the development of the technology.

The specific investment costs stated in (Görner and Lindenberger, 2015) are based on different literature sources. The costs for alkaline electrolysers (1,000 €/kW to 5,000 €/kW) show a very wide range due to scale effects. The specific costs of 1,000 €/kW refer to a plant on the MW scale. For a PEM electrolyser the specific investment costs in the year 2014 are significantly higher at about 2,000 €/kW.

In the article of (Antoni and Kostka, 2012) a price level of approximately 1,800 €/kW is given for an alkaline electrolysis plant in the lower MW range. The price is based on offers.

(Felgenhauer and Hamacher, 2015) analysed 16 offers of commercially available electrolysers in the power range of 0.35 to 3.35 MW. The investment costs include the electrolyser, transportation, installation and commissioning. The investment costs of AEC range from 2,100 \$/$kW_{H2\text{-}LHV}$ (output 54 kg_{H2}/h) to 5,700 \$/$kW_{H2\text{-}LHV}$ (output 5,9kg_{H2}/h). The PEMEC have higher costs between 3,100 \$/$kW_{H2\text{-}LHV}$ (output 47 kg_{H2}/h) and 6,600 \$/$kW_{H2\text{-}LHV}$ (output 9 kg_{H2}/h). This results in specific investment costs of approximately 875 €/kW_{el} (3.35 MW) to 2,370 €/kW_{el} (0.35 MW) for AEC and 1,370 €/kW_{el} (3 MW) to 2,915 €/kW_{el} (0,6 MW) for PEMEC.

The investment costs for a PEM electrolyser in (Walker et al., 2016) are derived from data provided by industrial partners. They are assumed to be about 1,500 $/kW (~ 1,130 €/kW) for a 5 MW electrolyser and about 1,250 $/kW (~940 €/kW) for a 30 MW electrolyser.

(Siemens, 2015) state that the second product generation of PEM electrolysers with a nominal power of 1.25 MW costs well below 2,000 €/kW where this used to be over 10,000 €/kW installed load. A further decrease down to even 900 €/kW was expected for the year 2018 at the latest.

According to information of the (Energieinstitut an der JKU Linz, 2018) from different projects, which are based on requests in the year 2017, the investment costs for alkaline electrolysers are in the range of about 900 – 2,500 €/kW at a power of about 0.5 – 2.5 MW. For PEM electrolysers the costs are about 1,600 – 2,000 €/kW for a rated power of 0.5 – 2 MW.

The share of investment costs of the electrolyser stack in the total electrolyser system costs is not often discussed. Apart from the earlier mentioned estimate of (Steinmüller et al., 2014) where the stack costs approximately 50% of the total, other estimates are given by (Hofstetter et al., 2014) and (Noack et al., 2014) where the stack is estimated to cost about 30% and 32% of the total investment costs, respectively.

Within STORE&GO, two demonstration sites use an alkaline electrolyser. The plant in Italy reports investment costs of roughly 1,400 €/kW (200 kW), which is thus in the same range as literature sources suggest. For the German demonstration plant investment costs of the electrolyser are currently unknown. The PEM electrolyser system in the Swiss demonstration site (350 kW, bought in 2014) was estimated to cost around 951,395 CHF (~€808,686), which equals to 2718 CHF/kW (~2311 €/kW). This price includes the electrolyser stack, rectifier, grid connection, control system, drying system, piping, measurement equipment and water treatment and the investment costs are indeed almost twice as high as those of an alkaline electrolyser.

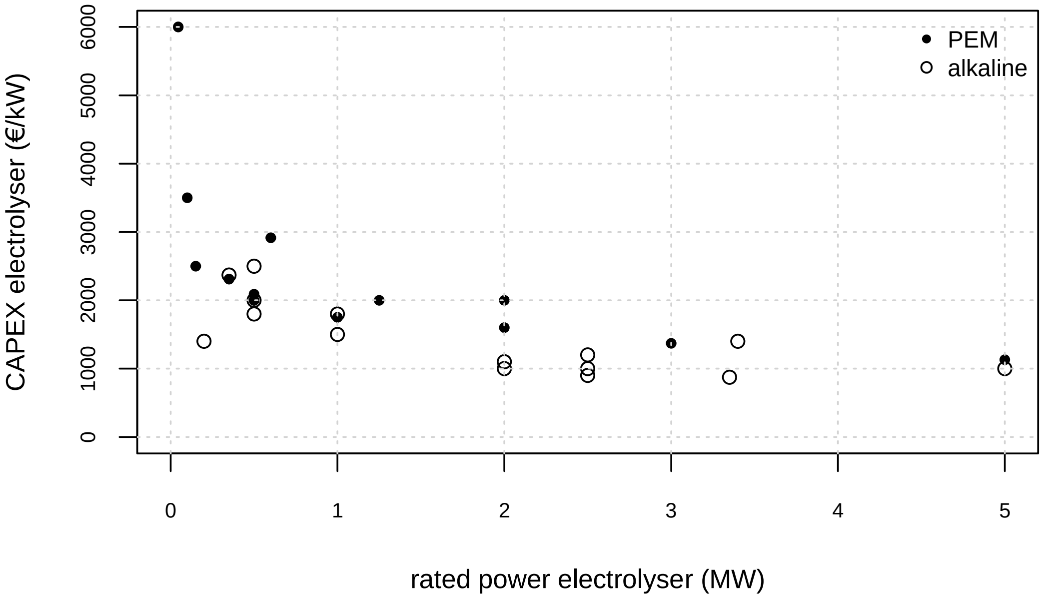

Figure 2: Overview of investment costs of alkaline and PEM electrolysers plotted against the reported rated power according to the literature sources discussed in this section. Sources: (Antoni and Kostka, 2012; Bertuccioli et al., 2014; de Bucy, 2016; Energieinstitut an der JKU Linz, 2018; Energinet dkEnergi Styrelsen, 2012; Felgenhauer and Hamacher, 2015; Görner and Lindenberger, 2015; Schenuit et al., 2016; Siemens, 2015; Smolinka et al., 2011; Steinmüller et al., 2014; Walker et al., 2016)

Figure 2 summarizes the cost estimates for PEM and alkaline electrolysers that were discussed in this section. The cost estimates are plotted against the rated nominal power to give an indication of the influence of size on the investment costs. In cases that the nominal power was only broadly indicated (e.g. “in the MW-scale”) a number has been chosen that seems to be representative. Two cost estimates (for rated powers of 10 and 30 MW respectively) are not visible in the figure.

As can be seen in Figure 2, investment costs of electrolysers tend to become lower with increasing size. The same conclusion was found by (Saba et al., 2018) who made an overview of (alkaline and PEM) electrolyser investment costs over the past 30 years. It was found that investment costs decrease significantly especially until 100 Nm^3/h (~0.5 MW electrical input) due to the peripheral costs which are independent of the plant size. The authors of the study also found that investment costs of electrolysers decreased significantly over the past decades, especially for PEM electrolysers of which the costs are now approaching those of alkaline electrolysers.

PEM electrolysers are relatively new and became only recently available in larger sizes (Saba et al., 2018). To properly compare investment costs of the two technologies, size needs to be taken into account. Table 1 presents averages for the investment costs calculated from the literature sources presented in this section that discuss electrolysers with rated powers between 1 and 5 MW and 0.5 and 5 MW respectively.

Table 1: Investment costs of alkaline and PEM electrolysers based on estimates from literature discussed in this section taking into account estimates of 1 – 5 MW and 0.5 – 5 MW.

		Alkaline		**PEM**
	Costs (€/kW_{el})	**Rated power (MW)**	**Costs (€/kW_{el})**	**Rated power (MW)**
1 – 5 MW	1180	2.5	1640	2.4
0.5 – 5 MW	1390	2.1	1890	1.6

Table 1 makes clear that even for similar rated powers, PEM electrolysers are currently more expensive than alkaline electrolysers. Nevertheless, the difference is smaller than it seems to be when the rated powers are not taken into account.

A more detailed investigation of investment costs of electrolysers, including learning effects and economies of scale will be given in STORE&GO deliverable 7.5[i].

Lifetime, efficiency and operational costs

The lifetime of the electrolyser depends on what efficiency drop is accepted. An electrolyser is unlikely to break down but reduces in efficiency until the point that investing in a new electrolyser stack becomes beneficial (Bertuccioli et al., 2014). (Smolinka et al., 2011) reports that alkaline electrolysers need revision every 7 – 12 years for some components but that other components last for 20 years and do not need replacement. The lifetime is reported to be 90,000 hours and no further improvements are expected. For PEM electrolysers the lifetime is reported to be only 20,000 hours but it is stated that the lifetime improved significantly over the last decade and further improvements up to 60,000 hours are expected for the next 10 – 20 years. (Bertuccioli et al., 2014) writes that leading alkaline and PEM manufacturers claim stack efficiencies of 90,000 and 60,000 operating hours respectively, meaning that the expected PEM lifetime improvements of (Smolinka et al., 2011) were already reached 3 instead of 10 – 20 years later. (Siemens, 2017) reports a lifetime of at least 80,000 hours for their SILYZER 200 PEM electrolyser, indicating an even further improvement. (Nel, 2018) states that cell stack replacement for their alkaline electrolysers is typically needed after 8 – 10 years.

Besides investment costs also the efficiency of an electrolyser is important, as this determines the required feedstock (electricity and water) needed to produce a certain amount of hydrogen and the hourly production volume of the plant, which is in turn important for the sizing of the hydrogen storage and methanation reactor. The efficiency is often defined in terms of electricity consumed per normal

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cubic meter (kWh/m^3, NTP) of produced hydrogen. To express the efficiency in terms of a percentage, the heating value (also known as calorific value or energy value) of hydrogen is required. The heating value of a substance is the amount of heat released during complete combustion and is different for every substance. There is a difference between the higher heating value (HHV) and lower heating value (LHV) of a substance. The HHV (also known as gross calorific value) includes the heat of vaporization of water whereas the LHV (also known as net calorific value) does not. The HHV and LHV of hydrogen are equal to 3.54 kWh/m^3 (NTP) and 3.00 kWh/m^3 (NTP) respectively.

In a perfect situation, 3.54 kWh of electricity is needed to create 1 m^3 H_2 (NTP). In reality, however, there is always a loss of energy in the electrolysis process so the electricity needed for 1 m^3 (NTP) will be higher. Larger units work more efficiency and so the energy consumption decreases with volume (Smolinka et al., 2011).

Similar to the determination of costs, it is important for the determination of the efficiency to specify whether the whole system is included or only the electrolyser stack. The efficiency of the bare electrolyser is higher than it is for the complete system. Furthermore, the efficiency of the electrolyser is also dependent on the operation mode and load: electrolysers are more efficient at lower loads (to a certain point) (Bertuccioli et al., 2014; Hydrogenics, 2017; Kopp et al., 2017).

(Frank et al., 2018) developed an extensive method to calculate the efficiency of PtG plants – including both electrolysis and methanation. The paper makes clear that for a proper comparison of efficiencies of different electrolysers and PtG plants, all system boundaries need to be identified carefully, just as the use of thermal energy. Unfortunately, current literature sources and electrolyser manufacturers generally do not clearly report the system boundaries of the efficiency calculations and the reported numbers are therefore hard to compare. Nevertheless, we will provide an overview of electrolyser efficiencies reported in literature.

(Gahleitner, 2013) made an evaluation of PtG pilot plants and also looked at the efficiencies of the plants. The average power consumption of alkaline and PEM electrolysers in the pilot plants was found to be 5.1 kWh/m^3 (69% efficiency) and 5.6 kWh/m^3 (63% efficiency) respectively. Lowest power consumptions were found to be 4.2 kWh/m^3 (84% efficiency) and 4.5 kWh/m^3 (79% efficiency) respectively for the two types of electrolysers. So, although literature suggests that PEM electrolysers can be more efficient than alkaline electrolysers, this was not found in practice in the pilot plants comparison of (Gahleitner, 2013). The author remarks, however, that the efficiencies of the pilot plants are hard to compare due to insufficient documentation about the efficiency calculation. Besides this, five years have passed since the article was published and PEM electrolysis has been further developed since then. (Kopp et al., 2017) presents more recent information on the operation of a PEM electrolyser and report efficiencies of 64% (HHV) for operation at part load and 59% (HHV) for operation at peak load. A PtG plant from Thüga in Frankfurt (Germany) that uses a PEM electrolyser reports a system efficiency of 77% (HHV) (ThügaMainova, 2017).

Several electrolyser manufacturers also report their power consumption / efficiencies. (Nel, 2018) reports power consumption of their alkaline electrolyser stack to fall in the range of 3.8 – 4.4 kWh/m^3 (93 – 80%). (Hydrogenics, 2018a) reports power consumption of their HySTAT alkaline electrolyser outdoor systems (all-in) to be 5.4 kWh/m^3 (66% efficiency) for small-scale electrolysers and 5.2 kWh/m^3 (68% efficiency) for large-scale electrolysers (>30 m^3/h) at full capacity. For indoor systems, power consumptions of 4.9 kWh/m^3 (72% efficiency) are reported. In another document (Hydrogenics, 2016) reports no difference in power consumption or response time between their HySTAT alkaline and HyLYZER PEM electrolysers, with both having a power consumption of 5.0 – 5.4 kWh/m^3 (71 – 66% efficiency) with utilities included at nominal capacity. (ProtonOnSite, 2017) reports system power consumption for their PEM electrolysers of 5.3 kWh/m^3 (67% efficiency). The 1.25 MW SILYZER PEM electrolyser stack of Siemens reports a production of 225 m^3/h, thus equalling to a power consumption of 5.6 kWh/m^3 (63% efficiency) (Siemens, 2017).

Several literature sources estimate the yearly operational costs of electrolysers. (Carr et al., 2014) use an estimate of 4% of investment costs for the yearly operational costs based on several literature sources without indicating a specific type of electrolyser. (Greiner et al., 2007) also uses operational costs of 4% of investment costs in their calculations, assuming an alkaline electrolyser is used. (Hofstetter et al., 2014) takes into account operational costs of 2% of investment costs based on several literature sources and assuming the use of an alkaline electrolyser. (Ulleberg et al., 2010) also assume that the yearly operational costs are 2% of investment costs for an alkaline electrolyser. (Bertuccioli et al., 2014) estimates the operational costs based on figures provided by manufacturers and found that these are in the range of 2 – 5% of investment costs per year, without a distinction between alkaline and PEM electrolysers. The authors state that operational costs differ by plant size and become lower for larger plants. For a smaller plant of about 1 MW they estimate operational costs of 5% of CAPEX per year while this would reduce to 2% for a 10 MW plant.

Summary

Table 2 summarizes the information presented in this section and gives a range and base case for the electrolyser CAPEX (both alkaline and PEM), lifetime, efficiency and OPEX.

Table 2: Base case assumptions and ranges for current costs parameters of electrolysers

	Alkaline		**PEM**	
	Base case	**Range**	**Base case**	**Range**
CAPEX (€/kWh)	1180	875 – 1800	1640	1130 - 2000
Lifetime	90,000 h (10 years)	-	60,000 h (7 years)	-
System efficiency (% HHV)	69%	66 – 72%	66%	59 – 77%
OPEX (% of CAPEX)	4%	2 – 5%	4%	2 – 5%

For the investment costs, average values are taken from Table 1 and the provided ranges represent the lowest and highest values found for electrolyser sizes between 1 and 5 MW. The system efficiencies reported in the table are calculated by averaging all values reported in this section for alkaline and PEM electrolyser systems respectively. As was said before, these values are difficult to compare because of unclear system boundaries. The model of (Frank et al., 2018) is recommended for future calculations and specification of the efficiency of PtG plants. This provides clear system boundaries and designations for unambiguous allocation and comparability of efficiencies. For the operational costs of the electrolyser, no distinction is made between alkaline and PEM electrolysers.

2.1.2 Hydrogen storage

The produced hydrogen can be stored before it is used, transported to another location, reconverted back to electricity or further converted into methane. The size of the storage facility strongly depends on the size of the plant but also on the application and the operation of the electrolyser. When the electrolyser is not operated continuously but instead is adapting to the fluctuating power supply of a wind farm or to fluctuating electricity prices, a storage facility is needed to buffer the fluctuating hydrogen production. An application with a non-continuous hydrogen demand – such as a hydrogen refuelling station – also requires a buffer storage. For seasonal hydrogen storage, very large storage volumes would be needed.

The technology PtG is currently still in development and existing plants are pilot plants of relatively small size. (Gahleitner, 2013) gives an overview of pilot plants and found that the power input of the electrolysers in the plants ranged between as little as 1 kW to a maximum of 6.3 MW for the plant of Audi that is located in Werlte, Germany. Another large (for current standards) PtG plant is located in Mainz, Germany, and has a PEM electrolyser with a 6 MW power input (Kopp et al., 2017). The three demonstration plants in the STORE&GO projects have electrolysers with a power input of 2 MW, 350 kW and 200 kW respectively. For onsite storage at this relatively small scale, high-pressure gas tanks and metallic hydride tanks are the best options (Götz et al., 2016). In the overview of pilot plants, given by (Gahleitner, 2013), most of the current projects (88%) and all of the planned projects use high-pressure gas tanks. This technology is state-of-the-art and is widely available with high capacities and relatively low costs (Gahleitner, 2013). (Zoulias et al., 2006) state that metal hydrides may compete with high-pressure storage tanks only for small storage capacities (some tens of m^3), even in the long-term. As hydrogen storage is a high cost factor, it is beneficial to minimise this as much as possible (Götz et al., 2016). In this report, we will focus on high-pressure steel tanks for small- to medium-scale hydrogen storage. Underground hydrogen storage in geological formations will be discussed for large-scale, long-term storage, which might become relevant in the future.

High-pressure steel tanks

The size of the storage facility depends strongly on the plant configuration and utilization of the hydrogen. The pressure of hydrogen storage varies between 4 and 400 or even 700 bars. The higher the storage pressure, the more gas can be stored on the same footprint, but the higher the costs will be due to the need for stronger storage tanks (thicker walls and/or other materials) and higher energy consumption of the compressor (Ulleberg et al., 2010). Using a high-pressure electrolyser can prevent the use of a compressor (Gahleitner, 2013) – or at least reduce the need for compression. Although this is more efficient, (Ulleberg et al., 2010) state that it results in higher costs for material, safety and control systems and therefore recommend low-pressure electrolysis followed by compression.

Estimates for investment costs in pressure tanks for hydrogen storage vary. Table 3 gives an overview of costs estimates for hydrogen storage in high-pressure tanks according to literature. All cost estimates have been translated into the same unit €/m^3 (NTP), which indicates the costs to store one normal cubic meter of hydrogen, irrespective of the pressure used in the tank.

(Carr et al., 2014) estimate the current costs of 200 bar hydrogen storage tanks at 500 €/kg (equal to 45 €/m^3). Three different future cost-scenarios are assumed, with hydrogen storage prices going down to 250, 125 or even only 50 €/kg (equal to 22, 11 and 4 €/m^3 respectively). They assume different storage sizes, equal to 100 h, 10 h and 1 h of full output of a wind farm.

(Darras et al., 2015) use a price of 70 €/m^3, without giving much information about the storage.

(Gammon et al., 2006) give costs for hydrogen storage at the HARI (UK) demonstration site where 48 cylinders of 0.475 m^3 are installed which can store a maximum of 2856 m^3 when the pressure is 137 bar.

(Grond et al., 2013) assumes capital costs for hydrogen storage of about 900 €/kg which equals to 81 €/m^3. The estimate was based on data from (Weinert, 2005), who gives a price range of 273 – 2182 $/kg (254 – 2029 €/kg or 23 – 182 €/m^3) based on literature sources. The literature sources report storage capacities between 50 – 1240 kg (556 – 13,793 m^3) with pressures ranging between 2057 – 8000 psi (141 – 552 bar).

(Karellas and Tzouganatos, 2014) assume capital costs of €422,700 for a storage tank of 1000 kg, which comes down to 38 €/m^3.

(Katikaneni et al., 2014) assesses the operation of on-site hydrogen production for a refuelling station. In a refuelling station, the hydrogen storage facility consists out of two parts: bulk storage at moderate pressure and a cascade storage at a high pressure. Here, costs for the 300 kg (3337 m^3) hydrogen bulk storage tank (173 bar) are estimated at 1200 $/kg (1140 €/kg or 102 €/m^3) and costs for the 140 kg (1557 m^3) cascade storage tank (432 bar) are estimated at 1150 $/kg (1093 €/kg or 98 €/m^3).

(Linnemann and Steinberger-Wilckens, 2007) calculate the costs of hydrogen production for two different plant sizes. They report amongst others the costs for 500 bar storage tanks for a hydrogen refuelling station (5000 m^3 in one example and 267 bundles of 3840 m^3 in a second example), which equals to 40 €/m^3.

(Ozaki et al., 2014) present a comparative study with costs estimates for a very large hydrogen storage based on 883 clusters containing 36 cylinders of 0.8 m^3 each. One cylinder of 0.8 m^3 can store up to 204 m^3 of hydrogen at a pressure of 350 bar. The costs for one cylinder are 1.0 million YPI (€8387).

(Pääkkönen et al., 2018) performs a techno-economic analysis of PtG in a biogas plant. Hydrogen is assumed to be stored at ambient pressure in an 85 m^3 steel tank, which is equivalent to 2-hours of operation of the anaerobic digester. Investment costs were determined using a formula from a chemical engineering handbook that bases investment costs on size. For the 85 m^3 tank they were found to be €42,000 which equals to 490 €/m^3.

(Prince-Richard et al., 2005) use a price of 200 $/kg (190 €/kg or 17 €/m^3), which was taken from a range of 50 – 500 $/kg (48 – 475 €/kg or 4 – 43 €/m^3).

(Ulleberg et al., 2010) give costs for the hydrogen storage at the Utsira (Norway) demonstration site where there is one storage tank of 12 m^3 that can store the hydrogen up to a pressure of 200 bar (2400 m^3).

(X. Xu et al., 2017) gives an overview of a hydrogen refuelling station with a storage system that consist of a bulk storage tank with a pressure of 173 bar followed by a cascade storage at a pressure of 350 bar – the pressure that is needed for dispensing. The costs for the storage tanks are calculated using a formula in which costs depend on the size and pressure of the tank. Costs range between 2300 and 1300 $/kg (2200 – 1200 €/kg or 195 – 110 €/m^3) for storage tanks of sizes between 14 and 300 kg of hydrogen. These estimates are very high compared to most other estimates presented here.

(Zoulias et al., 2006) gives an overview of hydrogen storage costs in high-pressure steel tanks based on cost data from commercially available storage tanks. For large storage volumes (2,000 – 10,000 m^3) at 200 bar, costs of roughly 30 – 40 €/m^3 were found. Smaller storage volumes showed in general higher prices, going up to 140 €/m^3. The authors estimate the current price at 38 €/m^3 and estimate a future price of 25 €/m^3 already for the year 2020.

The literature overview makes clear that cost estimates for hydrogen storage tanks vary widely. Prices generally fall in the range of 20 – 100 €/m^3, with extremes going down to only 4 €/m^3 for very optimistic future estimates and up to almost 500 €/m^3 for pessimistic estimates. Although (Zoulias et al., 2006) states that costs increase for smaller storage systems, this does not become apparent from the literature overview. Combining multiple gas cylinders usually creates larger storage volumes (e.g. (Gammon et al., 2006; Linnemann and Steinberger-Wilckens, 2007; Ozaki et al., 2014), which leaves not much room to reach significant economies of scale. (Simbeck and Chang, 2002) state that higher pressures also lead to higher costs but this also does not become apparent in the literature review. For the refuelling station studied by (Katikaneni et al., 2014), the high-pressure, smaller sized storage tank even has lower costs per m^3 than the bulk storage tank with a lower pressure and larger volume – although the price difference is minimal and could be caused by rounding errors.

In the STORE&GO demonstration sites, hydrogen storage is only needed for intermediate storage between the electrolyser and the methanation reactor. The plant in Falkenhagen (Germany) does not use intermediate hydrogen storage. In Troia (Italy) the original hydrogen plant has hydrogen storage but for the methanation process this is not needed. In Solothurn, Switzerland, hydrogen storage of 292 m^3 is used, at a maximum of 30 bar, which is a capacity of roughly 4.5 hours of full capacity operation. The containers can be used up to a pressure of 200 bar but they are filled to a maximum of 30 bar since this is the maximum outlet pressure of the electrolyser. The investment costs were estimated at 400,000 CHF (€340,000), which translates into 1370 CHF/m^3 (~1164 €/m^3), which is very high compared to literature estimates.

Table 3: Investment costs for hydrogen storage according to literature

Literature source	Pressure (bar)	Storage size	Costs (€/m^3)

(Carr et al., 2014)	200	Different sizes	45 → 22 → 11 → 4
(Darras et al., 2015)	Not specified	Not specified	70
(Gammon et al., 2006)	137	2856 m^3 (60 m^3 each)	50
(Grond et al., 2013)	Not specified	Not specified	81
(Karellas and Tzouganatos, 2014)	Not specified	11,123 m^3	38
(Katikaneni et al., 2014)	173	3337 m^3	102
(Katikaneni et al., 2014)	432	1557 m^3	98
(Linnemann and Steinberger-Wilckens, 2007)	500	5000 m^3	40
(Ozaki et al., 2014)	350	204 m^3	41
(Pääkkönen et al., 2018)	1	85 m^3	490
(Prince-Richard et al., 2005)	414	Different sizes	17 (4 – 43)
(Ulleberg et al., 2010)	200	2400 m^3	23
(Weinert, 2005)	141 – 552	556 – 13,793 m^3	23 – 182
(X. Xu et al., 2017)	350	3337 – 156 m^3	110 – 195
(Zoulias et al., 2006)	30	5 – 10,000 m^3	38 → 25

The lifetime of hydrogen storage tanks is assumed to be 20 to 30 years (Carr et al., 2014; Darras et al., 2015; Glockner et al., 2006; Greiner et al., 2007; Ramsden et al., 2008; Ulleberg et al., 2010; Zoulias et al., 2006), so no replacement during a project lasting for 20 years is required.

For the hydrogen storage, operating costs are estimated at 0.5% (Glockner et al., 2006; Karellas and Tzouganatos, 2014; Zoulias et al., 2006), 1% (Darras et al., 2015; Ramsden et al., 2008), 2% (Greiner et al., 2007) and 2.5% (Ulleberg et al., 2010) of initial CAPEX. In this report, we assume that the O&M costs of the hydrogen storage are equal to 1.5%, taking the average of the estimates from literature.

Table 4 gives the range and base case assumptions for the CAPEX, lifetime and OPEX of high-pressure steel tanks for hydrogen storage.

Table 4: Base case assumptions and ranges for current costs parameters of high-pressure steel tanks for hydrogen storage.

	Base case	Range
CAPEX hydrogen storage (€/m^3)	100	23 – 195
Lifetime	>20 years	20 – 30 years
OPEX (% of CAPEX)	1.5%	0.5 – 2.5%

Geological formations

There are several types of geological formations that can be suitable for long-term large-scale hydrogen storage. Geological formations are already widely used to store natural gas to bring flexibility into the gas system. Depleted gas and oil reservoirs, aquifers and salt caverns are the most important geological formations for underground natural gas storage (Ozarslan, 2012).

For underground hydrogen storage, salt caverns form the most suitable geological formation. Salt is inert with respect to hydrogen and it is extremely gas tight (Crotogino et al., 2010). This can be different in depleted oil and gas reservoirs and especially in aquifers where the hydrogen might react with the minerals and microorganisms in the reservoir (Amid et al., 2016; Crotogino et al., 2010).

Besides the advantages of being inert and leak tight, salt caverns are also characterised by very high withdrawal and injection rates and low cushion gas requirements (Kepplinger et al., 2011; Ozarslan, 2012). Cushion gas is the gas volume required in the storage to maintain an adequate storage pressure. Although this gas is present in the reservoir, it cannot be used. The cushion gas requirements are estimated to be roughly 30% of the total volume (Lord et al., 2014). In calculating investment costs for salt cavern storage, cushion gas costs must be included. These costs roughly fall in the range of 20 – 50 % of total investment costs (Le Duigou et al., 2017; Lord et al., 2014).

The technology of storing hydrogen in salt caverns is very similar to the storing of natural gas in these reservoirs. The energy density of hydrogen is however roughly three times lower than that of natural gas, which means that less hydrogen can be stored in the same volume, making hydrogen storage more expensive than the storage of natural gas. There are already several hydrogen storage facilities in underground salt caverns in the world: in Teesside (England) and in Texas (USA). Experience shows that the losses are negligible at less than 0.1% per year (Kepplinger et al., 2011).

The possibilities for geological storage of hydrogen are strongly dependent on the location. The distribution of salt formations across Europe is very uneven. The – by far – most favourable locations in Europe are in the north-west of Germany and north-east of the Netherlands (Crotogino et al., 2010) but these large salt deposits (Zechstein) are also spread to (east) England, Denmark and Poland (Kepplinger et al., 2011). Other areas in Europe also have salt deposits and although they are less favourable for hydrogen storage they might still be useful: e.g. (Le Duigou et al., 2017) identified several suitable locations for hydrogen storage in salt caverns in France.

Salt caverns for gas storage have typical volumes of 500,000 – 700,000 m^3 with hydrogen working gas capacities of roughly 8.5 kg hydrogen per m^3 (Crotogino et al., 2010; Kepplinger et al., 2011). The investment costs for salt caverns are highly dependent on the location (Le Duigou et al., 2017; Lord et al., 2014). In general, underground storage in geological formations is cheaper than aboveground storage but it is only suitable for large-scale applications (Schoenung, 2011). Several literature sources discuss the investment costs of hydrogen storage in salt caverns. (Schoenung, 2011) estimates gas storage costs for underground and aboveground at 15 and 0.3 \$/kWh respectively, with aboveground storage thus being 50 times more expensive than underground storage. (Le Duigou et al., 2017) estimate cavern investment costs at 9 €/m^3. (Steward et al., 2009) estimate the costs for the development of solution-mined salt caverns at 0.02 \$/kWh. Using the HHV of hydrogen of 39.41 kWh/kg and a cavern capacity of 8.5 kg/m^3 (following e.g. (Crotogino et al., 2010; Kepplinger et al., 2011) this equals to 6.7 \$/m^3 which is thus the same order of magnitude as the estimate of (Le Duigou et al., 2017). (Lord et al., 2014) gives an extensive overview of costs for geological storage of hydrogen in salt caverns, depleted oil and gas reservoirs, hard rock reservoirs and aquifers. Levelized costs for hydrogen storage in salt caverns were found to be 1.61 \$/kg, including cushion gas costs that added 18% to the total costs. Using again the cavern density of 8.5 kg/m^3 the costs are 13.7 \$/m^3, which is again in the same order of magnitude. Only the estimate of (Schoenung, 2011) is one order of magnitude higher than the others. Using only the three other estimates, we can estimate hydrogen storage costs in underground salt caverns at 10 €/m^3 which is equal to 0.03 €/kWh or 1.2 €/kg or 0.11 €/m^3.

The lifetime of salt caverns is long and estimated at 30 years (Lord et al., 2014) or even 50 years (Le Duigou et al., 2017). Replacement within a project lifetime of 20 years is therefore not required.

For hydrogen storage in salt caverns, O&M costs are estimated at 2% of investment costs per year (Le Duigou et al., 2017).

Table 5 gives the base case assumptions and ranges of the cost parameters of hydrogen storage in underground salt caverns.

Table 5: Base case assumptions and ranges for current costs parameters of hydrogen storage in salt caverns

	Base case	**Range**
CAPEX hydrogen storage (€/m^3)	0.11	0.06 – 0.85
Lifetime	>20 years	30 – 50 years

OPEX (% of CAPEX)	2%	-

2.1.3 Methanation reactor and BoP

Hydrogen produced in an electrolyser can be used directly in for example the chemical industry or transportation sector but it can also be further converted into methane. This conversion has several benefits since methane has many applications and our current economy is largely based on natural gas, which is a gas mixture of primarily methane. For conversion of hydrogen into methane a methanation reactor is required, just as a source of CO_2. There are two different methanation technologies: biological methanation and catalytic thermochemical methanation. Both technologies will be demonstrated within STORE&GO.

According to (J. Xu and Froment, 1989) there are 11 reactions that can take place in a methanation reactor, but three of those can describe the process:

$\mathbf{CO + 3H_2 \leftrightarrow CH_4 + H_2O}$ (R1) **Equation 2**

$\mathbf{CO_2 + H_2 \leftrightarrow CO + H_2O}$ (R2) **Equation 3**

$\mathbf{CO_2 + 4H_2 \leftrightarrow CH_4 + 2H_2O}$ (R3) **Equation 4**

These reactions are named the methanation reaction (R1), the reverse water-gas shift (RWGS) reaction (R2) and the Sabatier reaction (R3) (Zoss et al., 2016). Methanation is the reverse of methane steam reforming, which is currently the most important technique to produce hydrogen (Holladay et al., 2009).

(Götz et al., 2016) give an extensive overview of catalytic methanation. For this, a catalyst is required for the production of methane. Several metals are suitable to serve as a methanation catalyst but nickel is most often used due to the high activity and methane selectivity and low raw material price. Nickel requires, however, high purity of the gas that is fed into the reactor, mainly with regard to the hydrogen sulphide (H_2S) content, which can poison the catalyst (e.g. (Barbarossa and Vanga, 2011). There are several reactor types possible, with different research groups currently working on their development. The reactors are usually operated at temperatures in the range of 200 – 550 °C and the reaction is highly exothermic, making temperature control in the reactor an important issue.

Biological methanation is different from chemical methanation and uses microorganisms (archea) instead of a catalyst for the conversion of hydrogen and CO_2 into methane. It takes place under moderate conditions with low temperatures (20 – 70 °C) and ambient pressures. The main advantage of biological methanation over chemical methanation is that the archea can tolerate relatively high concentrations of impurities, removing the need for a very pure stream of CO_2 (Götz et al., 2016; Hofstetter et al., 2014). Nevertheless, after treatment might still be required depending on the further utilization.

Biological methanation can take place in a separate reactor but can also be done in situ in a biomass digester (Bensmann et al., 2014; Götz et al., 2016). In the latter option, hydrogen is fed directly into a biomass digester where biomasses are digested to form biogas – a gas consisting out of predominantly methane (roughly 60%) and CO_2 (roughly 40%). Biogas can be used directly but can also be upgraded to biomethane by removing the CO_2 and thereby upgrading the methane content of the gas, making injection into the natural gas grid possible. By adding hydrogen to the biomass digester, part of the CO_2 can be converted into methane. This methanation method removes the need for an additional methanation reactor but the conversion is limited to the CO_2 production rate of the biogas plant and total conversion of the produced CO_2 might be difficult (Götz et al., 2016).

In this report we only consider biological methanation in a separate reactor. In this method, any source of CO_2 can be used, including CO_2 from a biogas (upgrading) plant. Biological methanation is still in the development phase. (Götz et al., 2016) found that the supply of hydrogen to the microorganisms is the rate-limiting step in all reactor designs they investigated. Within STORE&GO, the Swiss demonstration site in Solothurn will be demonstrating the biological methanation.

Just as is the case with an electrolyser, a methanation reactor does not consist of a bare reactor only. Additional equipment is needed in the form of piping, measurement equipment, structure housing, gas purification, heating and cooling. The costs of these components – together forming the balance of plant (BoP) of the methanation reactor – need to be included in the final investment costs as well.

Investment costs

The methanation technology is, just as the electrolysis technology, currently still in the development phase and information on costs of methanation reactors is even more limited than it is for electrolysers. For this reason, the indication of investment costs for methanation plants are fraught with great deal of uncertainty (currently there are no cost data of commercial facilities in the context of PtG available). As the manufacturers mostly keep current specific costs for methanation plants in confidence, hardly any real costs could be determined. The reviewed literature sources provide therefore a rough estimate of the costs of a methanation reactor, for both chemical and biological methanation.

The determination of investment costs is further complicated by a variety of different processes, reactor types and operating modes. Furthermore, in most of the analysed studies, the system boundaries of the indicated investment costs are not well defined, thus further limiting the comparability. Usually the specific investment costs are given in €/kW$_{CH4}$ (rated CH_4 output power). If the costs are related to the rated power of the electrolyser of the PtG plant the unit is €/kW$_{el}$. The two can be converted into each other using the combined efficiency of the electrolysis and methanation process.

Chemical methanation

Two – rather old – literature sources report costs for methanation of biomass. (Gassner and Maréchal, 2009) investigate the impact of process pressure for thermochemical production of SNG from lignocellulosic biomass. For the methanation taking place at 15 bar specific costs of about 190 €/kW$_{SNG}$ are expected compared to approx. 550 €/kW$_{SNG}$ at 1 bar. (Zwart et al., 2006) estimate the specific investment costs of a chemical methanation reactor at about 580 €/kW$_{SNG}$ for a 10 MW$_{th}$ bio-SNG plant (production of SNG from biomass) run at atmospheric pressure. For a larger plant with a power of 100 MW$_{th}$, which is run at a pressure of 7 bar, the costs drop to about 107 €/kW$_{SNG}$

More recent literature sources report costs for methanation reactors in PtG plants.

In a techno-economic study of PtG concepts, (Graf et al., 2014a) estimate the total investment costs (apparatus, steel construction, foundations, electrics, instrumentation, and engineering) of catalytic methanation plants for three different sizes (5 MW$_{SNG}$, 30 MW$_{SNG}$ and 110 MW$_{SNG}$). Depending on the size, the total investment costs for the plants are €1.5 million, €4.9 million or €12.1 million, which leads to specific investment costs of around 300 €/kW$_{SNG}$, 160 €/kW$_{SNG}$ or 110 €/kW$_{SNG}$.

In a PtG system analysis, (Steinmüller et al., 2014) estimates investment costs for the methanation part of a PtG plant (electrical input 48 MW) to be about 140 €/kW$_{el}$ (~250 €/kW$_{SNG}$).

(Grond et al., 2013) provide a graph, based on several sources, for investment costs of chemical methanation plants against the rated power. Costs are decreasing with increasing capacity and are about 1,500 €/kW$_{CH4}$, 1,000 €/kW$_{CH4}$ and 750 €/kW$_{CH4}$ for sizes of 1 MW$_{CH4}$, 3 MW$_{CH4}$ and 6 MW$_{CH4}$ respectively. This result in costs related to the input power of the electrolyser of about 840 €/kW$_{el}$, 560 €/kW$_{el}$ and 420 €/kW$_{el}$ (by assuming a combined efficiency of 56 %). However, it is mentioned, that due to the fact that currently no small methanation (< 20 MW$_{CH4}$) plants are offered as a standard or mass-produced product on the market, the investment costs seem to be relatively high. The costs are expected to drop to 300 – 500 €/kW$_{CH4}$ (170 – 280 €/kW$_{el}$), if the market for small-scale methanation develops.

For an assessment of different PtG process chains, (Schiebahn et al., 2015) estimate the investment costs of a chemical methanation plant. Costs are estimated at 720 €/kW$_{CH4}$, which is equal to approximately 400 €/kW$_{el}$ by assuming a combined efficiency of 56%.

(Götz et al., 2016) carried out a literature review on PtG. The specific investment costs reported in the study have a wide range from 130 – 1,500 €/kW$_{SNG}$ and are not very reliable. It is stated that the cost estimates from Outotec GmbH (see also (Graf et al., 2014a)) based on size might be most realistic: 400 €/kW$_{SNG}$ for a 5 MW and 130 €/kW$_{SNG}$ for a 110 MW plant. Other costs estimates are therefore probably too high.

(Albrecht et al., 2013) analyse the costs of producing renewable gases. Specific investment costs for a 3.7 MW$_{CH4}$ methanation plant are estimated for the current year (2013) as well as for the future. For 2013 and 2016 investment costs were estimated to be ~3,300 €/kW$_{CH4}$ and ~2,000 €/kW$_{CH4}$ respectively, which is much higher than other estimates. It is noted, however, that the assumptions are rather conservative and that further cost reductions are conceivable but cannot be assessed on the basis of current data.

Two of the three STORE&GO demonstration plants use chemical methanation. The plant in Falkenhagen (Germany) uses a two stage chemical methanation. The first stage is a honeycomb and a wall coated reactor in parallel, but only one is used at the same time. The second stage is a fixed bed reactor for polishing the gas quality. All three catalytic methanation reactors use nickel as a catalyst. The costs for the plant are estimated at roughly €1.7 million including all equipment but excluding costs for construction, preparation, planning, installation and design. These costs are estimated at roughly €1.4 million, which is an addition of about 80% to the basic plant components costs. Especially costs for civil works are very high (~€600,000), which is due to current capacity issues in the construction industry. Potential costs savings of the complete plant are estimated at a maximum of 30%. The costs for the plant in Troia (Italy) are not known yet but an estimate can be given based on the grant proposal where planned costs were ~€0.9 million, including all components but excluding costs for installation planning and design. The plant in Falkenhagen is thus estimated to be much more expensive as the one in Troia but the plant is five times larger (1 MW[j] versus 0.2 MW for the Italian plant), making the relative costs much lower. Related to the electrical input of the electrolyser, methanation reactor costs for the German and Italian demonstration plants (including only components) are equal to 1,700 and 4,400 €/kW$_{el}$ respectively, or 3,000 and 7,900 €/kW$_{CH4}$ (assuming a combined efficiency of 56%). The difference in costs confirms the effect of scale that was also found in literature. A detailed evaluation of the STORE&GO demonstration plants (including a cost evaluation) will be given in Deliverable 5.9[k].

Biological methanation

In (Graf et al., 2014b) an assessment of biological methanation is performed. Specific investment costs for methanation reactors (including engineering, construction, machinery and peripherals but without the production and provision of hydrogen) are estimated based on information of two manufacturers and found to be in the range of 340 – 1200 €/kW$_{SNG}$ depending on the size (110 – 1 MW) and configuration of the plant. Out of all the investment costs, 65 – 80% is for the machinery, 8 – 18% for the construction, around 10% for planning and design and 4 – 8% for other things.

(Krassowski, 2012) mentions costs of ~1,150 €/kW$_{SNG}$ for rather small methanation plants (130 kW$_{SNG}$ output) and ~100 €/kW$_{SNG}$ for larger plants (10 MW$_{SNG}$ output), indicating that scale has a very large impact.

(Petersen, 2016) estimates the investment costs of biological methanation at 400 €/kW$_{el}$ for a 2 MW$_{el}$ PtG plant in the year 2017 (~715 €/kW$_{SNG}$). At the beginning of the year 2016 costs were twice as high.

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[j] Falkenhagen has a 2 MW electrolyser but the methanation part has only capacity for hydrogen of 1 MW$_{el}$ equivalent.

[k] D5.9: Final report on evaluation of technologies and processes. Due date: 29 February 2020.

In a factsheet for biological methanation plants, (BVES, 2016) ranges the current specific investment costs from 700 to 1,500 €/kW$_{CH4}$ for the bioreactor, engineering, approval and installation.

(Grond et al., 2013) also provide a graph for investment costs of biological methanation against the capacity (it is mentioned, that the data are from (Krassowski, 2012) but they differ from each another). The specific investment costs are decreasing with an increasing methane output of the plant and are about 320 €/kW$_{CH4}$, 120 €/kW$_{CH4}$ and 90 €/kW$_{CH4}$ for an output power of 200 kW, 1,000 kW and 2,000 kW respectively.

(Heller, 2017) quantifies the specific investment costs for a biological methanation demo plant (combination with an electrolyser with a rated power of 1 MW) in the year 2016 with 1,200 €/kW$_{el}$ (~1,800 €/kW$_{CH4}$).

(Budny et al., 2015) gives estimates for the total investment costs for a 5 MW PtG plant for hydrogen and methane, using biological methanation. For the PtG system, the estimate for the hydrogen plant is 600 €/kW$_{el}$ and for the methane plant 875 €/kW$_{el}$, meaning that the (biological) methanation part costs 275 €/kW$_{el}$ (~490 €/kW$_{SNG}$).

Based on an analysis on a 1 MW biological methanation plant for a wastewater treatment plant done by the company Electrochaea, (Hofstetter et al., 2014) assumes investment costs of 145 CHF/kW$_{el}$ for the methanation reactor and another 337 CHF/kW$_{el}$ for the BoP of the methanation reactor, in total 482 CHF/kW$_{el}$. This results in total specific investment costs for the methanation plant of about 690 €/kW$_{CH4}$ (overall efficiency 0.56; 1 EUR$_{2014}$ = 1.2 CHF$_{2014}$).

One of the three demonstration plants in the STORE&GO project uses biological methanation. In this plant, located in Solothurn (Switzerland), the methanation takes place in a stirred vessel. The CAPEX of the complete system was estimated at ~€1.4 million in the grant proposal, which translates into ~2,000 €/kW$_{el}$[l] or ~3,500 €/kW$_{CH4}$ (assuming a combined efficiency of 56%[m]). These costs are much higher than other reported costs. The reason for this is the cost intensive equipment that is used for evaluation of the technology within STORE&GO. A detailed evaluation of the STORE&GO demonstration plants (including a cost evaluation) will be given in deliverable 5.9[n].

Lifetime, efficiency and operational costs

None of the above sources provides an estimate of the lifetime of the methanation reactor. Within the STORE&GO consortium, it is expected that the reactor could last for 20 years although replacement of the catalyst (for chemical methanation) and maintenance is required, the frequency of which depending on the operation mode and plant design.

It is important to specify the definition of efficiency of the methanation reactor. There is a difference between conversion rate and energetic efficiency of the reaction, whereby the latter can also include the use of heat that is produced by the methanation reactor, improving the overall efficiency. For calculation of the efficiency, the chemical energy of the final product should be considered – following the methodology presented in STORE&GO deliverable 5.2 or (Frank et al., 2018). The efficiency of the methanation reactor is then defined as the energy content of the final product to the energy content of the hydrogen. For calculating this, the higher heating value (HHV) of hydrogen and methane should be used, which are 141.8 and 55.5 MJ/kg respectively.

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[l] The plant in Solothurn has an electrolyser output of 60 Nm^3/h but the methanation unit has an input of 120 Nm^3/h.

[m] First simulations in Solothurn showed a combined efficiency of about 40.5%, if a membrane is used to reach more than 96 mol-% of CH_4 and less than 2 mol-% H_2 without circulation of the permeate, which would give costs of ~7050 €/kW$_{SNG}$. In Switzerland a limited injection with less strict requirements of the gas quality is possible and the efficiency is higher than 56%.

[n] D5.9: Final report on evaluation of technologies and processes. Due date: 29 February 2020.

Table 6 shows the Sabatier reaction, translated into weight (kg) and energy content (MJ) using the molar masses and HHV of the substances for 100% and also 80% conversion.

At 100% conversion, 1.0 kg of hydrogen can be converted into 2.0 kg of methane. Using the HHVs of the two substances this means that 1 MJ of hydrogen is converted in 0.7785 MJ of methane. This means that the energetic efficiency is limited to 77.85% at 100% conversion. At a conversion rate of 80%, 1 MJ of hydrogen is converted into 0.6228 MJ of methane, thus indicating an efficiency of 62.28%. Since there is still 20% of the hydrogen left (thus equal to 0.2 MJ), however, the overall energetic efficiency becomes 82.28%, when you take into account all final products. The lower the conversion rate, the higher the energetic efficiency: when the methanation reactor does not convert hydrogen at all (conversion rate is 0%), the energetic efficiency is 100%.

Table 6: Sabatier reaction, weights and energy content (based on HHV) of the different substances

Sabatier reaction (100%)	$4H_2$	+	CO_2	→	CH_4	+	$2H_2O$		
kg (100% conversion)	1.0	+	5.5	→	2.0	+	4.5		
MJ (100% conversion)	1.0				0.7785				
Sabatier reaction (80%)	$5H_2$	+	CO_2	→	CH_4	+	$2H_2O$	+	H_2
kg (80% conversion)	1.0	+	4.4	→	1.6	+	3.6	+	0.2
MJ (80% conversion)	1.0				0.6228				0.2

Several literature sources report efficiencies of methanation reactors but it is not always clear what is meant exactly. Many literature sources refer to other sources that in turn refer to rather old sources that discuss the conversion of biomass into synthetic natural gas (SNG). For example, (Lehner et al., 2014) assume an efficiency of 80% for the methanation reaction. They refer to (Grond et al., 2013) where it is stated that the efficiency of chemical methanation is in the range of 70 – 85% based on sources that refer to the conversion of biomass into SNG. It is mentioned that the remaining energy is released as high temperature heat. For biological methanation, (Grond et al., 2013) assume a very high efficiency of 95-100% based on personal communication with the company Electrochaea. Because the energetic efficiency cannot be this high, it is likely that the terms efficiency and conversion rate are confused here. Biological methanation is unlikely to be more efficient than chemical methanation. (Schiebahn et al., 2015) also assume an efficiency of 80% for the methanation reaction and state that due to the exothermic character of the Sabatier reaction, about 17% of the energy is released as heat, thereby limiting the maximum efficiency at 83%. Both (Lehner et al., 2014) and (Schiebahn et al., 2015) assume an efficiency of 70% for the electrolyser, leading to a combined efficiency of 56% for the PtG plant. (Budny et al., 2015) assume an efficiency of 82% for the electrolysis and 78% for the methanation process, resulting in a combined efficiency of 64%. The Audi plant in Werlte reports a combined efficiency of 54% for the electrolysis and methanation process (Rieke, 2013).

(Hofstetter et al., 2014) define the efficiency as the ratio of output energy (HHV of the gas produced) to the input energy and states that for the methanation reaction the output gas is methane and the input energy is the hydrogen (HHV). This definition would imply that the energy content of potential hydrogen that is left is not taken into account in calculating the efficiency but the system-level efficiency range of 77 – 82% that they report for biological methanation (based on literature and expert interviews) is higher than the theoretical maximum of 77.9%. (Parra and Patel, 2016) report the efficiency of chemical methanation to be in the range of 70 – 85% based on literature sources whereby the efficiency is defined as the ratio of the calorific value of the output and input gases.

At the moment of writing, the STORE&GO demonstration sites are not in operation yet and real efficiencies of the methanation reactors cannot be given. The performance of the demonstration sites will be evaluated later in the project, mainly in deliverable 5.9[o]. Due to the confusing aspect of the efficiency of the methanation reaction, it can be clearer to use the conversion rate as a cost parameter in business model calculations.

Unfortunately, not many literature sources report the operational costs of methanation reactors. For biological methanation, (Hofstetter et al., 2014) assume operational costs to be equal to 5% of investment costs per year. Costs for nutrients are assumed to be negligible. (Budny et al., 2015) reports operational costs for a PtG plant (5 MW electrical input) producing hydrogen and a plant with an additional biological methanation reactor. The difference in fixed yearly operation costs between the two plants refers to the biological methanation part of the plant and comes down to 4% of investment costs of the methanation reactor. (Grond et al., 2013) estimate operational costs for chemical methanation to be 10% of investment costs per year, including replacement costs of the catalyst. Several literature sources that discuss the economics of chemical methanation refer to this source for the operational costs (e.g. (Chiuta et al., 2016; Gutiérrez-Martín and Rodriguez-Anton, 2016). (Graf et al., 2014b) do not give specific numbers for the operational costs but calculate the SNG generation costs in biological and catalytic methanation plants. It is found that these costs are in a very similar range. In-situ biological methanation was found to have slight advantages over catalytic methanation, while external biological methanation is more expensive. For large plant sizes (110 MW) catalytic methanation was found to be more profitable than biological methanation. (Giglio et al., 2015b) give a very detailed evaluation of operational costs for a PtG plant that integrates SOEC with catalytic methanation, thereby distinguishing between fixed operational costs (in €/y) and variable operational costs (in €/h). The costs often refer to the whole plant and cannot easily be separated for the methanation part only. For the STORE&GO demonstration plants, operational costs are not known yet.

Summary

Table 7 summarizes the ranges and base case assumptions of the cost parameters of methanation reactors (both chemical and biological methanation), including CAPEX, OPEX, lifetime and energetic efficiencies. For the CAPEX, ranges in the table are from the literature overview presented in this section. The base case estimates are based on the average and median of the values presented in this section (excluding the extremes). Investment costs for biological methanation are found to be slightly higher than those of chemical methanation, but this might be caused by the fact that this technology is less developed and average biological methanation plants are currently significantly smaller than chemical methanation plants. Overall, it can be concluded that the investment costs for methanation reactors are very uncertain and the ranges found are very large. The STORE&GO demonstration plant in Falkenhagen (Germany) reports costs to be in the upper range of the estimates presented in the table (1500 €/kW$_{SNG}$) and this still excludes costs for installation, preparation, construction, planning and design. In this report, those costs are taken into account separately, as will be discussed later in section 2.2. Just as for the electrolysers, the scale of the plant is important, and larger plants have lower investment costs. A more extended review of costs of methanation reactors – including economies of scale and learning curves – will be given later in the STORE&GO project in Deliverable 7.5[p]. The energetic efficiency is reported in Table 7 as 77.9% - indicating a conversion rate of 100%. The operational costs reported for chemical methanation are based on just one source and are therefore not very reliable. Also for biological methanation the operational costs are not very certain – being based on just two literature sources.

Table 7: Base case assumptions and ranges for costs parameters of methanation reactors.

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[o] D5.9: Final report on evaluation of technologies and processes. Due date: 29 February 2020.

[p] D7.5: Report on experience curves and economies of scale. Due date: 31 October 2018

	Chemical methanation		**Biological methanation**	
	Base case	**Range**	**Base case**	**Range**
CAPEX methanation reactor (€/kW_{SNG})	400	110 – 1500 (3300)	550	100 – 1500 (4300)
Lifetime (years)	20	-	20	-
Energetic efficiency (% HHV)	77.9%	-	77.9%	-
OPEX (% of CAPEX)	10%	-	5%	4 – 5%

2.1.4 CO_2 storage tank and compressor

To convert hydrogen into methane, a source of CO_2 is required. To guarantee a steady supply of CO_2 to the reactor, a storage facility needs to be installed. The size of the storage facility depends on the source of CO_2, the distance of the CO_2 source to the PtG plant and how it is delivered to the PtG plant. The CO_2 is usually stored in liquid form in a storage tank.

In the STORE&GO demonstration plant in Falkenhagen (Germany), a 30 m^3 storage tank for liquefied CO_2 is installed to ensure an uninterrupted supply of CO_2 to the plant. The tank stores CO_2 at a pressure of 16 – 20 bar at –24 °C and can ensure 11 days of full load operation of the plant without refilling. At Falkenhagen, CO_2 is delivered by truck. The tank is rented for a price of 42,000 €/year.

In the STORE&GO demonstration plant of Solothurn (Switzerland), a buffer tank of 2 m^3 (16 bar) is installed for storage of CO_2. Investment costs are reported to be 20,000 CHF (~€17,000). Besides the CO_2 storage tank itself there is also a compressor and pipeline needed. Costs of these are reported to be 120,000 CHF (~€102,000) and 70,000 CHF (~€59,500) respectively, making the total investment costs for CO_2 storage 210,000 CHF (€178,500). The compressor needs service every 2000 operation hours, costs of which are reported to be €1,500. At full-time operation, four services are needed every year, leading to costs of €6,000 which is ~6% of CAPEX of the compressor.

In the STORE&GO demonstration site in Troia (Italy) the CO_2 is delivered by an air capture plant developed by (Climeworks, 2018). The produced CO_2 will be compressed and stored in buffer tank. Investment costs for a gas balloon, compressor and buffer tank are €55,000. Operational costs are not known yet.

Table 8 summarizes the costs of CO_2 storage in the three STORE&GO demonstration sites. To be able to compare the costs better, the yearly costs have been translated into the present worth (PW) costs during 20 years, using a discount rate *i* of 6%. Furthermore, costs have been translated into the costs per MW installed electrolyser capacity to be able to compare the plants better. Costs for the plant in Troia are higher in practice, since yearly operational costs are not included here.

Table 8: Costs for CO_2 storage in the STORE&GO demonstration plants. Yearly costs have been translated into the costs during 20 years, using the present worth (PW) and a discount rate of 6%.

	Falkenhagen	**Solothurn**	**Troia**
Plant size (MW)	1 (2)[q]	0.7 (0.35)[r]	0.2
Investment costs	None	€178,500	€55,000

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[q] Falkenhagen has a 2 MW electrolyser but the methanation part has only capacity for hydrogen of 1 MW_{el} equivalent. In the calculations for total costs per MW installed electrolyser capacity we therefore use 1 MW instead of 2 MW.

[r] In Solothurn the methanation reactor has an input flow of 120 Nm^3/h whereas the electrolyser output is 60 Nm^3/h. In the calculations for total costs per MW installed electrolyser capacity we therefore use 0.7 MW instead of 0.35 MW.

Yearly costs	€42,000	€6,000	Unknown
Total costs (PW, 20 years)	~€482,000	~€247,000	>€55,000
Total costs per MW (PW, 20 years)	~€482,000	~€353,000	>€275,000

2.1.5 Compressors for hydrogen and methane

A compressor is needed to inject hydrogen, methane or CO_2 in a storage tank. Whether or not a compressor is required for injection of hydrogen or methane in the natural gas grid depends on the final pressure of the produced gas and the pressure in the grid. Different compressors are needed for different situations, and the design and costs depend on the pressure difference that needs to be overcome, the flow rate used and the type of gas that is compressed. In PtG plants, compressors might be needed for hydrogen, methane and CO_2. Compressors for CO_2 were shortly discussed together with CO_2 storage tanks in the previous section. Here, hydrogen compressors will be discussed first, followed by some remarks on methane compressors.

(André et al., 2014) provides a formula for the compression power P in kW for hydrogen compressors:

$$\boldsymbol{P = Q * \frac{ZTR}{M_{H_2} * \eta_{comp}} * \frac{N\gamma}{\gamma - 1} * \left[\left(\frac{P_{out}}{P_{in}} \right)^{\frac{\gamma-1}{N\gamma}} - 1 \right]} \qquad \textbf{Equation 5}$$

With in this formula:

- Q the flow rate in kg s^{-1}
- Z the hydrogen compressibility factor: set at 1 as an approximation
- T the temperature at the inlet of the compressor: set at 278 K
- R the ideal gas constant: equal to 8.314 J K^{-1} mol^{-1}
- M_{H2} the molecular mass of hydrogen: equal to 2.016 g mol^{-1}
- η the compressor efficiency: chosen as 75%
- N the number of compressor stages
- γ the diatomic constant factor: equal to 1.4
- P_{in} the inlet pressure of the compressor
- P_{out} the outlet pressure of the compressor

With this formula, the required power of the compressor can be calculated based on the flow rate and pressure difference that needs to be overcome. The maximum flow rate depends on the plant size, whereas the pressure difference depends on the destination of the gas: a pipeline or storage tank at a certain pressure.

(André et al., 2014) also give a formula to calculate the energy consumption of a compressor:

$$\boldsymbol{Energy\ consumption\ compressor\ (MWh) = \frac{hours/year}{DTE} \times P} \qquad \textbf{Equation 6}$$

With P the power of the compressor in MW and DTE the Driver Thermal Efficiency, which is put at 90%. The energy consumption of the compressor in MWh can be multiplied with the electricity costs to find the energy costs of the compressor.

Estimates for investment costs of compressors vary widely in literature. The price can be expressed in €/kW (electrical input). Several literature sources report costs for hydrogen compressors in PtG applications. (Gammon et al., 2006) report costs for a 3.75 kW compressor with a flow rate of 11 m^3/h. They are estimated at £59,000 (€69,400), which translates into 18,500 €/kW. (Ozaki et al., 2014) report costs for a 700 kW compressor with a flow-rate of 275 kg/h (3059 m^3/h). Total investment costs consist not only of the costs for the compressor (827 million JPY) but also of costs for coolers (100 million JPY) in a cooling tower (30 million JPY) including construction work (6 million JPY), leading to a total of 963 million JPY (€8,100,000), which translates into 11,570 €/kW. (Ulleberg et al., 2010) report a significantly lower price for a hydrogen compressor of 5000 €/kW.

(Hofstetter et al., 2014) assumes a compressor is needed to inject the generated hydrogen in a high-pressure transmission line. The investment costs for the compressor are estimated at 134 CHF/kW, which equals to roughly 144 €/kW. This amount is two orders of magnitude lower than the high estimates given by (Gammon et al., 2006; Ozaki et al., 2014). Transmission lines are operated in the pressure range of 50 – 70 bar, which is lower than the pressures in the storage tanks. Still, it appears to be a very low estimate of the price. (Ozaki et al., 2014) also discusses low-pressure (≈ 8.5 bar) storage of hydrogen. The costs of the compressor belonging to this low-pressure storage are estimated at €636,350 for a 3060 m^3/h flow and compressor of 100 kW. This results into an investment of 6365 €/kW, which is considerably lower than the high-pressure compressor (about half) in the same study but still much higher than the estimate of (Hofstetter et al., 2014).

(Greiner et al., 2007) also give a relatively low estimate for hydrogen compressor costs of 700 €/kW. (X. Xu et al., 2017) estimates capital costs for a hydrogen compressor at 1000 \$/kW (950 €/kW) and (André et al., 2014) estimate the costs of the compressor to be 2545 \$/kW (2435 €/kW).

(Yang and Ogden, 2007) give a formula for the CAPEX of the compressor, which is lower at higher kW:

$$\boldsymbol{CAPEX\ compressor}\ (\$) = \mathbf{15,000} * \left(\frac{\boldsymbol{S_x}}{\mathbf{10}\ \boldsymbol{kW}}\right)^{\mathbf{0.9}} \qquad \textbf{Equation 7}$$

To illustrate this formula: CAPEX costs are about 1800 €/kW to 1000 €/kW for compressors of 1 to 350 kW respectively.

The above overview illustrates that it is difficult to determine a proper estimate for the compressor investment costs. Estimated prices range from as low as 144 €/kW to as high as 18,500 €/kW. The higher the flow rate and pressure difference that needs to be overcome, the higher the costs.

The lifetime of the compressor is estimated at 10 (Greiner et al., 2007; Ramsden et al., 2008) or 12 years (Ulleberg et al., 2010) – making replacement within a project lifetime of 20 years necessary. Operating and maintenance costs for the compressor are estimated at 4% (Greiner et al., 2007; Grond et al., 2013; Ramsden et al., 2008; Yang and Ogden, 2007), 3% (André et al., 2014) or 1.5% (Ulleberg et al., 2010) of CAPEX.

In the STORE&GO demonstration plants, the use of compressors is limited. The plants in Falkenhagen (Germany) and Troia (Italy) do not use intermediate hydrogen storage and thus also do not need a hydrogen compressor here. The plant in Solothurn (Switzerland) uses intermediate hydrogen storage but the storage tanks are filled to a maximum of 30 bar, which is the outlet pressure of the electrolyser. A hydrogen compressor is thus not used here. The plant in Falkenhagen (Germany) was originally producing hydrogen that was injected into the natural gas grid. The pressure in the grid that is used for injection of the produced gas is 48 bar, meaning that a compressor is necessary. Now that the plant is extended with a methanation reactor, the existing hydrogen compressors will still be used. If a compressor were designed specifically for methane, however, it would have been significantly smaller. Investment costs for the compressor from 10 to 55 bar are €600,000. In Solothurn, the grid pressure is only 5 bar. Since the operation pressure is 10 bar, a compressor for grid injection of the produced methane is not needed here. In Troia (Italy) the produced methane is not injected into the gas grid but instead is liquefied and transported by trucks. Here, a compressor is needed for the liquefaction.

Table 9: Base case assumptions and ranges for costs parameters of hydrogen compressors

	Base case	**Range**
CAPEX hydrogen compressor (€/kW)	-	144 – 18,500
Lifetime	10 years	10 – 12 years
OPEX (% of CAPEX)	3%	1.5 – 4%

Table 9 gives an overview of base case assumptions and ranges of the compressor cost parameters. The overview focuses on hydrogen compressors. Because the range in investment costs is very wide and the design of the compressor is very dependent on the situation, no base case value can be given.

2.1.6 Pipelines

The costs for a pipeline depend not only on the distance that needs to be bridged, but also on the pressure, flow of the gas, surroundings that need to be crossed and type of gas that needs to be transported. Because of this, it is important to differentiate between hydrogen and methane, transmission (high-pressure) and distribution (low-pressure) pipelines and rural and urban areas. Several estimates from literature are discussed and compared in Figure 3.

(Butenko et al., 2012) give estimates for methane pipeline costs of 65,000 and 400,000 €/km respectively for distribution and transmission grids. The pipelines for transmission grids are more expensive due to the higher pressures and as a consequence higher requirements for the pipelines.

(André et al., 2014) make a comparison between trucks and pipelines to transport hydrogen from large-scale production facilities to refuelling stations. A formula is given to determine the investment costs of hydrogen pipelines based on the diameter (D) of the pipeline in mm:

$$\boldsymbol{CAPEX\ (\$/km) = 418869 + 762.8 * D + 2306 * D^2} \qquad \textbf{Equation 8}$$

(Castello et al., 2005) mentions that estimates for hydrogen pipelines vary widely (ranges between 600,000 and 1,000,000 $/km) and are contradictory. They propose to take the capital costs of natural gas pipelines as a reference and include extra costs for hydrogen, which would be improvement of welding, joining, lining, coating, etc. The correction factor would range between 1.4 and 2.0 depending on the pipeline diameter and technology. Costs for methane and hydrogen transmission pipelines are represented by equations 6 and 7 respectively, with in these formula D the diameter of the pipelines in meters.

$$\boldsymbol{CAPEX\ (M\$/km) = 1.989 * D} \qquad \textbf{Equation 9}$$

$$\boldsymbol{CAPEX\ (M\$/km) = 4.10 * D - 2.11 * D^2} \qquad \textbf{Equation 10}$$

According to (Castello et al., 2005), different rules apply for distribution pipelines. Costs for the construction in heavily urbanized areas are very high compared to the pure material costs, making these pipeline costs not very dependent on pipeline length and diameter and which gas is transported. Costs were determined to be 100 and 500 €/m for areas with low or high urbanization respectively. The high price of 500 €/m was determined based on the replacement of piping in a trafficked road in the centre of a large European city, for which costs include identifying of other pipelines, cutting the existing road surface, divert traffic, etc.

(Yang and Ogden, 2007) divide the costs for hydrogen pipelines in material costs and costs for installation, right-of-way and miscellaneous. The latter can vary greatly depending on the location. The installation costs are estimated at 300,000 and 600,000 $/km respectively for transmission (rural) pipelines and distribution (urban) pipelines. Transmission and distribution pipelines are classified as such only by their location (rural or urban), and not by pressure requirements. To the construction costs, material costs are added, which are calculated based on the diameter (D) in inches according to:

$$\boldsymbol{Pipeline\ capital\ costs\ (\$/km) = 1869 * D^2} \qquad \textbf{Equation 11}$$

Figure 3 presents the discussed cost estimates for hydrogen and methane transmission and distribution pipelines in both rural and urban areas. Note that in this overview, transmission pipelines are cheaper than distribution pipelines according to (Yang and Ogden, 2007) because they are easier to install as they are in rural areas, whereas according to (Butenko et al., 2012) transmission pipelines are more expensive due to higher pressures and thus higher requirements and costs.

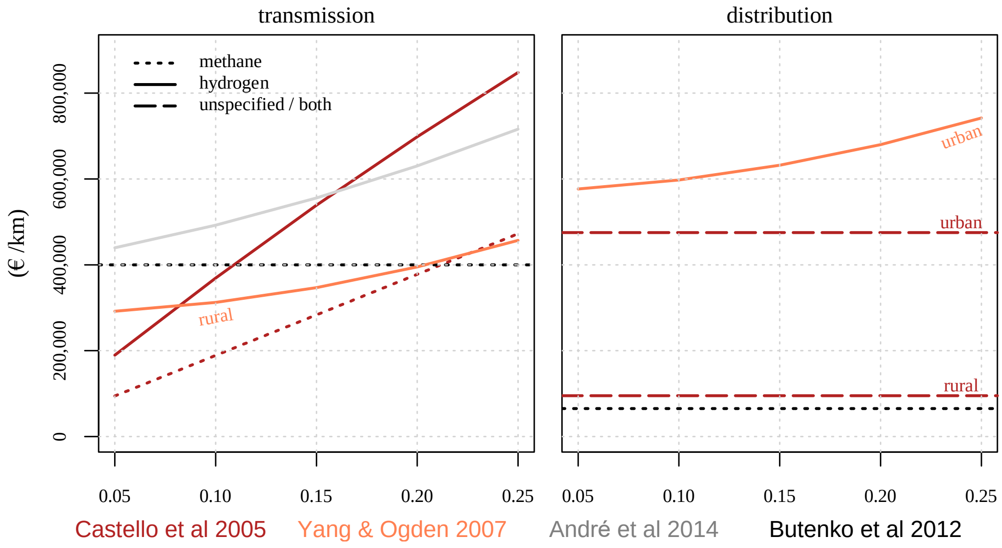

Figure 3: Cost comparison of pipelines transporting methane and hydrogen from four different literature sources (André et al., 2014; Butenko et al., 2012; Castello et al., 2005; Yang and Ogden, 2007) with on the x-axis the diameter in meters. The left and right graphs show cost estimates for transmission and distribution pipelines respectively. Dotted lines, normal lines and dashed lines show costs for methane, hydrogen and unspecified (both) pipelines respectively. Estimates provided in dollars were converted into euros by using an exchange rate of $1 = €0.95.

Based on the overview presented in Figure 3, several conclusions can be drawn. First, there is a large difference in costs of a pipeline depending on the location. In highly urbanized areas, a large part of the costs for a pipeline do not arise from the pipeline materials, but for the construction itself. Second, transmission pipelines are in general more expensive than distribution pipelines, due to the higher pressures and, as a consequence, higher specifications for the materials. Third, hydrogen pipelines are more expensive than pipelines for methane transportation, due to required improved joining, welding and coating. Costs are about 1.4 to 2.0 times as high, according to (Castello et al., 2005). For distribution pipelines, the differences between methane and hydrogen are smaller – as the costs for digging and construction are by far the most important contributors here to the final costs.

One can conclude that transmission pipelines are expensive because of the high requirements needed because of the high pressures, and distribution pipelines are much cheaper when looking at material costs, but due to the fact that most of them are in highly populated areas, costs can be very high and comparable to the transmission pipelines. Table 10 presents estimates for the different types of pipelines based on the literature sources that were discussed here.

(Schoots et al., 2011) analysed investment costs of onshore transmission pipelines (30 cm diameter) for methane, hydrogen and CO_2. It was found that investment costs for the pipelines vary strongly and are very dependent on the circumstances such as the country and region were they are built. Average costs were found to be 715,000, 788,000 and 854,000 $/km (~679,000, 749,000 and 811,000 €/km) for methane, CO_2 and hydrogen pipelines respectively, including costs for materials, labor, right-of-way and miscellaneous. The bandwidths of the cost estimates are very large (one order of magnitude) and labor costs represent roughly half of the total costs. The costs fall in the range of the costs presented in Figure 3.Transporting CO_2 differs from transporting methane because at the general transportation pressure and temperature levels, CO_2 is a supercritical fluid that requires booster stations with fluid pumps. The construction material can be the same as for methane pipelines: as long as the relative humidity stays low enough corrosion can be prevented. Hydrogen pipelines need different materials because hydrogen can diffuse into steel and cause hydrogen embrittlement and hydrogen attack. Transporting hydrogen at low pressures and temperatures avoids material corrosion. For higher energy flows more robust materials are needed which would lead to higher costs for both construction and operation and maintenance. The authors did not observe any significant cost reductions throughout time and also do not expect serious cost reductions in the near future.

In the STORE&GO demonstration sites, there are no very long pipelines for hydrogen or methane transport present. At the demonstration site in Troia (Italy), the produced gas will be liquefied on site. In Falkenhagen (Germany) and Solothurn (Switzerland) the produced methane will be transported over a short distance to be injected in the local gas grid. The plant in Solothurn reports investment costs of a 100m methane pipeline to be 10,000 CHF (€8,500) and a 55m hydrogen pipeline to be 5,500 CHF (€4,675). Both pipelines have thus equal costs of 85,000 €/km, which is rather low compared to the literature estimates.

The lifetime of pipelines is estimated to be very long: (Qadrdan et al., 2015) for example estimates lifetime for transmission and distribution pipelines to be 60 and 80 years respectively. Operational costs for pipelines are estimated at 2% of investment costs (André et al., 2014; Butenko et al., 2012).

Table 10: Cost estimates for transmission and (rural and urban) distribution pipelines for the transportation of hydrogen and methane, based on the estimates summarized in Figure 3.

	Transmission	**Distribution**		
		Rural	**Medium**	**Urban**
CAPEX methane pipeline (€/km)	350,000	100,000	300,000	500,000
CAPEX hydrogen pipeline (€/km)	600,000	350,000	450,000	550,000
Lifetime	60 – 80 years			
OPEX (% of CAPEX)	2%			

2.1.7 Gas grid injection station

Hydrogen produced in a PtG plant can be injected into the natural gas grid. It does, however, influence the combustion behaviour of the gas in the grid, and can therefore only be admixed up to a certain limit to prevent serious effects on material integrity. The allowable limit is location-specific and depends not only on the gas grid infrastructure but also on the specific end-users (Grond et al., 2013; Müller-Syring et al., 2013). Every country has set its own, often conservative, limitations. In the Netherlands, for example, the current admixture limit is only 0.02% (de Joode, 2014; Grond et al., 2013) and it is planned to be increased up to 0.5% (Verhagen, 2012). In Germany, the hydrogen admixture limit is an order of magnitude higher. It is currently 5.0% (Schiebahn et al., 2015) and it is planned to be increased up to 10.0% (Grond et al., 2013) but only at locations that do not have technical limitations or safety issues. The two STORE&GO demonstration plants in Falkenhagen (Germany) and Solothurn (Switzerland) both report a maximum hydrogen limit of 2%. The admixture limit determines together with the location-specific grid characteristics (pipeline flow, pressure, diameter, connected producers and end-users) the amount of hydrogen that can be injected.

Another important effect of hydrogen admixture in the natural gas grid is the changing of the energy density of the gas in the grid. The volumetric energy density of hydrogen is one third of that of natural gas (13 vs. 39 MJ/m^3) and thus a higher flow rate is required to meet the demand (Qadrdan et al., 2015).

Methane can be injected into the natural gas grid without the limitations that hydrogen has. The produced gas needs to fulfil certain quality requirements, however, before it can be injected, and the injection can still be restricted due to pipeline limitations. A limited gas demand during off-peak hours can be a bottleneck for the maximum amount of methane that can be injected continuously per hour.

(Holstein et al., 2011) give costs for a gas grid injection station, including gas quality measurement equipment. For the gas quality measurement equipment, investment costs of €50,000 are assumed, with operational costs of €10,000 per year. An injection point in the transmission grid is more expensive than an injection point in the distribution grid: the investment costs are estimated at €350,000 and €25,000 respectively.

(Albrecht et al., 2013) reports costs for methane gas grid injection stations in Germany and mentions that here, 75% of the grid connection costs must be borne by the grid operator. The plant has to pay the remaining part of the grid connection costs with a maximum of €250,000. For gas production of 116 m^3/h and 465 m^3/h (~ 2.2 and ~8.9 MW electrical input at an electrolyser efficiency of 69% and 100% conversion in the methanation reactor) the investment costs in a 45 bar grid are estimated to be €79,979 and €250,000 respectively – the latter thus equal to the maximum.

The STORE&GO demonstration plant in Falkenhagen (Germany) injects the produced methane in a transmission grid at a pressure of 55 bar. Investment costs are reported to be roughly €900,000 – including gas quality measurement and all other piping and instrumentation, but excluding a compressor to compress the gas from 10 to 55 bar. The costs represent the total costs without differentiating between costs made by the plant and gas grid operator.

The STORE&GO demonstration plant in Solothurn (Switzerland) reports gas grid injection investment costs of 35,000 CHF (€29,750) and 50,000 CHF (€42,500) respectively for costs of the injection station and quality measurement. Additional costs for a pipeline (10,000 CHF) were presented already in the previous section. Together, investment costs for the gas grid injection station equal to 85,000 CHF (€72,250). At this site, the produced gas is injected in a 5 bar grid.

(Holstein et al., 2011) assumes that the operational costs of a gas grid injection station are 5% and 2% of the investment costs for the transmission grid and distribution grid respectively[s]. The STORE&GO demonstration plant in Solothurn reports annual operational costs of 2000 CHF (~€1700), which equals to 2% of investment costs. (Albrecht et al., 2013) states that there are no operational costs for the PtG plant since the gas grid operator is responsible for this and bears the costs.

CAPEX and OPEX for the gas grid injection station vary with the chosen location, boundary conditions, hourly gas production and pressure level in the grid. Table 11 presents the estimates for the investment and operational costs of the gas grid injection station, distinguishing between the distribution grid (low pressure) and transmission grid (high pressure). For the investment costs of the transmission grid injection station, the maximum costs borne by the PtG plant in Germany are assumed here. Real costs can be much higher.

Table 11: Base case assumptions and ranges for costs parameters of gas grid injection stations

	Base case	Range
CAPEX gas grid injection station (distribution) (€)	75,000	72,250 – 75,000
CAPEX gas grid injection station (transmission) (€)	250,000	80,000 – 900,000
Lifetime (years)	NA	NA
OPEX distribution (% of CAPEX)	2%	-
OPEX transmission (% of CAPEX)	5%	-

2.1.8 Refuelling station

It is expected that, once fuel cell electric vehicles (FCEVs) running on hydrogen acquire a significant market share, hydrogen from a PtG plant could best be sold in refuelling stations. High prices are

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[s] For the transmission injection point, (Holstein et al., 2011) give the operational costs as a fixed number. This is converted into a percentage to make it more comparable to the operational costs of the distribution grid.

expected for renewable hydrogen at refuelling stations, compared to prices for hydrogen that could be obtained elsewhere (e.g. (Breyer et al., 2015; Grueger et al., 2017). Also methane produced in a PtG plant can be sold as a fuel in the transportation sector. When the produced methane can be labelled green[t], car owners can fuel their cars with green methane.

The concurrent design of electrolysers and hydrogen refuelling stations has been discussed in literature (e.g. (Prince-Richard et al., 2005; Symes et al., 2012; Zhao and Brouwer, 2015). Different companies are already offering and building self-contained modules for hydrogen refuelling that consist of an electrolyser, compressor, storage tank and dispenser (Hydrogenics, 2018b; ITM Power, 2018).

Depending on the circumstances, investments have to be done to be able to sell hydrogen to car owners. Assuming an already existing station, only a storage facility for the hydrogen – usually consisting of bulk storage at moderate pressure followed by a high-pressure cascade storage for the actual filling (Katikaneni et al., 2014; Ramsden et al., 2008; X. Xu et al., 2017) – and one or more dispensers are needed.

(Ramsden et al., 2008) estimates the price for two hydrogen dispensers in a refuelling station at $44,800, which is thus $22,400 (€21,280) per dispenser. (Katikaneni et al., 2014; X. Xu et al., 2017) both estimate the price of a dispenser at $30,000 (€28,500). The first assumes two dispensers are required at a refuelling station, while the latter assumes the number of dispensers to be dependent on the size of the fuel station. The number of dispensers is estimated to be 1, 2 and 4 dispensers for 10, 1000 and 2000 kg/day stations.

(Ramsden et al., 2008) estimate the lifetime of the dispenser at 10 years. They estimated operating and maintenance costs of dispensers to be 800 $/year for every dispenser, which equals to €1520 per year for two dispensers. There are also labour costs and other costs for the refuelling station, but when it is assumed that an existing refuelling station is extended with a hydrogen-filling device extra costs for personnel will be limited.

2.1.9 Oxygen and heat capture

In the electrolysis process, electricity is used to split water into hydrogen and oxygen. Although the goal of a PtG plant is to produce hydrogen, oxygen is also produced in large quantities (0.5 m^3 per m^3 of hydrogen or 8 kg for every kg of hydrogen) and it could be sold as well, thereby potentially improving the overall efficiency and economics of the plant. Nevertheless, (Graf et al., 2014a) reports that all PtG plants currently vent their produced oxygen to the atmosphere. Within the STORE&GO project, also none of the demonstration plants will capture the oxygen. The potential of selling oxygen is considered to be low (Graf et al., 2014a), but there might be some exceptions, mainly in cases where the produced oxygen could be used on site (E&E Consultant, 2014). The location of the PtG plant is very important for the potential selling of oxygen (Hofstetter et al., 2014) just as the balance between production and demand. If oxygen demand is not very large relative to the supply of the PtG plant, large quantities of oxygen will be wasted (Kato et al., 2005) and capturing it might not be economically feasible. To be able to sell the produced oxygen, it must be captured. Not many literature sources report costs for capturing the oxygen. (Hofstetter et al., 2014) reports investment costs for oxygen capture to be 24,100 CHF (€20,485).

Heat is another by-product of a PtG plant. The heat released from the electrolyser or a biological methanation reactor is relatively low-temperature heat. All STORE&GO demonstration plants report that the temperature of the released heat is too low to make it useable. Heat from a chemical

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[t] The issue whether or not methane produced in a PtG plant is “green” or not depends on both the source of the electricity in the electrolyser as well as on the source of CO_2 that is used. This topic is addressed already in STORE&GO deliverable 8.1: Exploring the future for green gases. Due date: 31 Augustus 2017.

methanation reactor can be of quite high-temperature. The potential for selling heat is strongly dependent on the location, as heat cannot be transported over long distances without considerable loss of energy. When heat can be used, investment and operational costs for coolers is reduced, just as the costs heating elsewhere.

2.2 Total investment and yearly operational costs of PtG plants

Besides the general investment costs for all separate components, extra costs have to be added for installation, land preparation, project planning, design, engineering, civil and site, control and safety, etc. Costs for these things are usually expressed as a percentage of total investment costs. Some literature sources that mention the investment costs for electrolysers specifically mention that these additional costs for installation, planning and design are already included (e.g. (de Bucy, 2016). For most literature sources, however, this is unclear. In this report, costs are taken into account for preparation, planning, design and installation. It is assumed these costs are 28% of the total investment costs for the equipment of the plant – following (Leonzio, 2017; Ramsden et al., 2008). Real costs can be higher, as is reported for example by the STORE&GO demonstration plant in Falkenhagen (Germany) for the methanation reactor. Here, additional costs for planning, preparation, construction, installation and design were 80% of the investment costs of the methanation plant components. The current capacity issues in the construction industry and the fact that the methanation reactors are relatively new cause these high costs, among others.

Table 12 gives an overview of the investment costs for a PtG plant producing methane using biological or chemical methanation. For simplicity, numbers are given for a plant with a 1 MW electrolyser, but the reported costs are valid for PtG plants with 1 – 5 MW electrolyser capacity and scale linearly with size (i.e. a PtG plant with a 2 MW electrolyser is twice as expensive as the plant presented in the table). For all investment costs, the base case assumptions presented earlier in this chapter are taken.

For the electrolyser, system costs of 1180 €/kW are taken into account, assuming an alkaline electrolyser is used as this is currently the cheapest type of electrolyser. We assume that the hydrogen storage tank is relatively small and can buffer 4 hours of full-load electrolyser production, similar to what is installed in the STORE&GO demonstration plant in Troia (Italy). For a 1 MW electrolyser with an efficiency of 69%, this means the storage tank must be able to store 780 m^3 of hydrogen. Investment costs are assumed to be 100 €/m^3 for storage in steel tanks[u], which means total investment costs for the hydrogen storage are ~€78,000. It is assumed that a hydrogen compressor is not required. For the CO_2 storage costs estimates from the Solothurn (Switzerland) plant are followed[v], scaled to fit a 1 MW plant: €255,000 for a storage tank, compressor and pipeline. For the methanation reactor we assume investment costs of 400 and 550 €/kW_{SNG} respectively for chemical and biological methanation. Assuming efficiencies of 69% for the electrolyser and 77.9% for the methanation (100% conversion), which is a combined efficiency of 53.8%, this translates into investment costs of ~215 and ~296 €/kW_{el} respectively for chemical and biological methanation. In the table, the produced methane is assumed to be injected in a local distribution grid, with investment costs of €75,000, including the injection point and measurement equipment. No compressor is needed for this injection in the grid.

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[u] Note that a typical salt cavern has a size of 500,000 – 700,000 m^3 and a working gas capacity of 8.5 kg/m^3 (see section 2.1.2.2). This would mean a typical salt cavern (500,000 m^3) could hold 4.25 million kg of hydrogen, which is equal to over 47 million Nm^3 of hydrogen, which is a completely different order of magnitude than the storage of only 4 hours of production mentioned here.

[v] The reason fort his is the fact that this STORE&GO demonstration plant is the only plant for which both investment costs and operational costs are currently available.

Table 12: Current investment costs for a PtG plant producing methane, based on a 1 MW electrolyser

	Component	Chemical methanation	Biological methanation
1	Electrolyser system (alkaline)	€1,180,000	€1,180,000
2	Hydrogen storage	€78,000	€78,000
3	CO_2 storage facility	€255,000	€255,000
4	Methanation reactor (system)	€215,000	€296,000
5	Gas grid injection station	€75,000	€75,000
Total components		€1,803,000	€1,884,000
Installation, planning, design (28%)		€504,840	€527,520
Total (million €)		€2.31 million	€2.41 million

Table 12 makes clear that the electrolyser is by far the most important component of a PtG plant in terms of costs. The investment costs are roughly half of the total investment costs of the plant for both the chemical and biological methanation plant. The methanation reactor adds about 9 – 12% to the total costs, depending on the technology used. The investment costs of the hydrogen storage facility are very limited here, but it has to be noted that the assumed size (4 hours of storage capacity) is quite low – especially for a PtG plant that would adapt to fluctuating electricity prices or fluctuating power output from e.g. a wind farm. In addition, extra costs for hydrogen pipelines and compressors are not included. In many plant configurations the costs for hydrogen storage are therefore likely to be higher.

The fixed operational and maintenance costs of a PtG plant consist amongst others out of administration, insurance and inspection and maintenance of the equipment including spare parts and personnel costs. The costs are made independent of the amount of hours the plant is in operation (although more maintenance might be required when the equipment is more intensively used). The costs do not include end-of-life replacement and are defined as a % of CAPEX for the different components that were specified earlier in this chapter. For the CO_2 storage facility we follow the costs specified by the demonstration site in Solothurn (Switzerland) that has operational costs of roughly 6000 €/year which is ~3.5% of CAPEX per year. As was said before, operational costs for the methanation reactors are based on just two sources: one for each technology and are therefore not very reliable. The fact that operational costs for the chemical methanation plant are higher than those of the biological methanation plant is based on just these two sources.

Table 13 shows the operation and maintenance costs of the different types of PtG plants.

Table 13: Operational and maintenance costs for a PtG plant with biological and chemical methanation. The costs are defined as a percentage of the investment costs, where the investment costs are taken from Table 12.

Component	% of CAPEX	Chemical methanation	% of CAPEX	Biological methanation
Electrolyser system	4%	€47,200	4%	€47,200
Hydrogen storage	1.5%	€1,170	1.5%	€1,170
CO_2 storage facility	3.5%	€8,925	3.5%	€8,925
Methanation reactor (system)	10%	€21,500	5%	€14,800
Gas grid injection station	2%	€1,500	2%	€1,500
Total yearly O&M costs (€)		~€80,000		~€74,000

All components with a lifetime shorter than 20 years need replacement within a project lifetime. These include the electrolyser and the compressors, which both need replacement after roughly 10 years. For calculations of e.g. hydrogen or methane production costs including all costs over the whole lifetime of a plant, these replacement costs need to be incorporated. This is not included yet in the operational costs presented in Table 13.

3 Feedstock costs of PtG plants

3.1 Water

Water is relatively cheap and its costs do not add much to the total costs of operating an electrolyser. In the Netherlands, water prices are based on a fixed tariff for the water per m^3, a tariff for the water connection per year, depending on the size of the connection and taxes for the first 300 m^3 of water that is delivered (Vitens, 2017). The fixed tariff in 2017 at the company Vitens (delivering water to a large part of the Netherlands) is equal to 0.668 €/m^3 (including VAT). For a PtG plant, the additional charge for the connection and the taxes for the first 300 m^3 do not add much to this number, due to the large amounts of water consumed in the plant. Because of that, it is possible to assume a fixed price of 0.69 €/m^3 (0.00069 €/kg) for further business model calculations, no matter the scale and operating mode of the plant.[w]

The actual requirement of water is more than the theoretical amount based on stoichiometrics. This would be roughly 0.8 L/m^3 hydrogen whereas electrolyser manufacturers report tap water consumptions in the range of 0.9 – 2.0 L/m^3 hydrogen (Hydrogenics, 2016; Nel, 2018; Siemens, 2017). Assuming the need for water to be 200% of the stoichiometric need (1.6 L/m^3), 17.9 kg water is needed to produce 1 kg of hydrogen, which means water costs are 0.01233 €/kg hydrogen.

To calculate the water costs for methane production, the efficiency or conversion rate of the methanation reaction is required, in addition to the information already given. Assuming a conversion rate of 100% (77.9% efficiency), 2 kg of methane can be produced from 1 kg of hydrogen (see also Table 6), which means that 9 kg of water is needed to produce 1 kg of methane. Costs of water are 0.00620 €/kg methane.

3.2 Electricity

Electricity is the most important feedstock of a PtG plant. Together with the investment costs, it basically determines the hydrogen production costs.

3.2.1 Electricity markets

Power producers and consumers can plan their production and consumption in electricity markets. Electricity can be traded years before actual delivery in forward markets but also close to real time in short-term electricity markets. The most important electricity market in terms of traded volume and number of market participants is the day-ahead (DA) market. In the DA market, market participants place bids for every hour with the volume and price of the electricity they want to sell or buy. Based on all the bids, computer algorithms on exchanges determine a market price for every hour[x]. The DA market closes one day before delivery. After closure of the DA market, the intraday (ID) market opens. In the ID market, power producers and consumers can adapt their market position using new information that became available after closure of the DA market. Examples of this are new weather forecasts – changing the expected wind or solar production – or information about plant outages. The ID market closes just before actual delivery starts: 60, 30, 20 or 5 minutes before, depending on the country. After closure of the ID market, the transmission grid operator (TSO) becomes responsible for balancing the electricity grid. Power production and consumption need to be in balance all the time and market participants have the responsibility to maintain this balance and stick to the scheduled production or consumption. Nevertheless, deviations from the schedule will occur – if only

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[w] In other countries, costs for water can be significantly higher than they are in the Netherlands. In Switzerland, for example, costs for water are generally more than 2 CHF/m^3 (1.7 €/m^3) (Schweizerische Eidgenossenschaft, 2018)

[x] In addition to trade at exchange platforms, there is also Over-the-Counter trading where brokers set the price.

because the weather and demand cannot be predicted with 100% accuracy – and TSOs need to be able to activate reserve capacity.

The provision of these reserve capacities is arranged on forehand in the balancing market. Power producers and consumers can offer (part of) their capacity to this balancing market. When necessary, they will be activated. Most electricity markets have three types of control reserve capacities, varying in their activation time and duration: primary control, secondary control and minute (tertiary) control. Primary control needs to be activated immediately as soon as deviations in the grid occur. Power producers and consumers offering primary control need to be able to offer both positive and negative control – i.e. they need to be able to both decrease and increase their power production or consumption. The power plants receive a fee for offering the reserve capacity. In secondary and minute control, power producers and consumers can choose whether they offer negative control (needed when too much electricity is fed into the grid, thus requiring increased consumption or decreased production) or positive control (needed when production is lower than consumption, thus requiring decreased consumption or increased production). In these reserve capacity markets, power consumers and producers receive a fee when they are actually used for balancing the grid, in addition to the fee for offering the reserve capacity.

PtG plants can buy electricity in long-term forward markets or in short-term DA or ID markets. They could also offer (part of) their capacity to the balancing market. The latter is discussed in several literature sources. Surprisingly, these sources are not conclusive on which market (primary, secondary, minute, positive or negative control) is the best option for a PtG plant to participate. (Guinot et al., 2015) and (Breyer et al., 2015) both state that a PtG plant could best participate in the primary reserve market, while others (e.g. (Baumann et al., 2013; Grueger et al., 2017; Kopp et al., 2017; Kroniger and Madlener, 2014; Lück et al., 2017) state that this is not an option and propose participating in the secondary or minute control reserve market. The choice for the different control reserve markets seems to be mainly influenced by the assumed technical limitations of the electrolyser: is it fast enough to deal with the prerequisites of the primary control reserve. The fact that plants offering primary control need to be able to both deliver negative and positive control is also a reason why some literature sources (Baumann et al., 2013; Kopp et al., 2017) state that primary control is not an option as it would require the electrolyser to run at base load, of which costs would be too high and exceed possible revenues of the primary reserve capacity price. For the same reason, these authors also only consider negative (secondary and minute) control. (Lück et al., 2017), on the other hand, found that positive control was more profitable than negative control due to the operation scheme. Operation in the DA spot market, however, was found to be most profitable.

In this report, we will further use recent DA electricity prices in Germany and the Netherlands.

3.2.2 DA electricity prices

When a PtG plant is operated continuously throughout the year, the costs for electricity are based on the average electricity price. Figure 4 shows recent average DA electricity prices in Germany and the Netherlands[y]. The increase observed in Germany in 2017 was not limited to Germany. Also other countries such as France and Denmark showed higher average DA electricity prices in 2017 than in earlier years (Bloomberg LP, 2017; Nord Pool, 2017).

It is important to note that costs for electricity are not limited to the DA price for most power consumers. There are additional costs in the form of fees and taxes. Germany, for example, has the EEG surcharge: an amount of money consumers have to pay on top of the electricity price for stimulating renewable energy in Germany. Since it was introduced, the EEG surcharge is increasing every year and nowadays (since 2013) it is higher than the average DA market price, reaching 68.80 €/MWh in

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[y] For the Netherlands, no data are currently available for 2017

2017. Some large power consumers are exempted from paying the EEG surcharge, but at the moment existing PtG pilot plants are not. For further evaluation of the PtG business case in this report we only take into account the market prices (DA) of electricity, but it is important to keep in mind that costs are usually higher in reality.

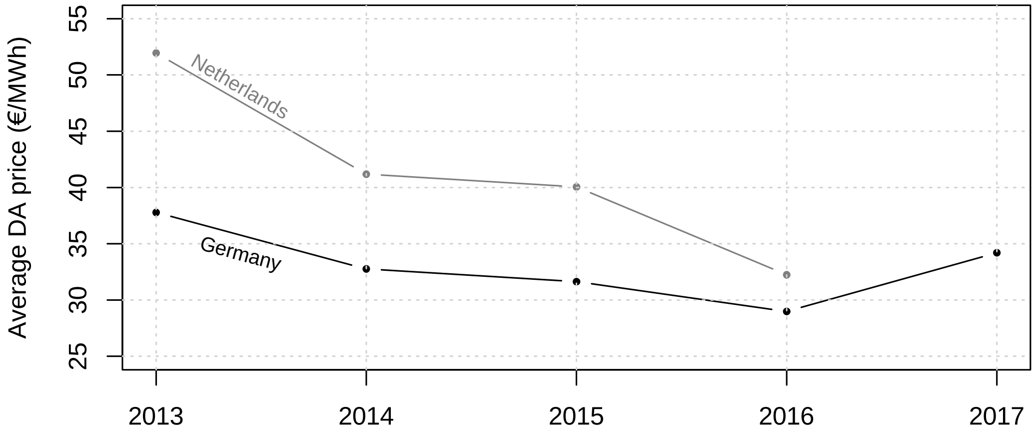

Figure 4: Recent average DA electricity prices in Germany and the Netherlands. Source: (Bloomberg LP, 2017)

It is often stated that PtG can be used to absorb surpluses of renewable energy in times of high wind or solar production and that a PtG plant can profit from low and even negative electricity prices. To assess the potential of PtG, it is interesting to know how often these low electricity prices occur and whether the electricity price patterns vary (strongly) among countries and years. Figure 5 shows electricity price patterns (average price for the X% cheapest hours in a year) for Germany and the Netherlands for the different years.

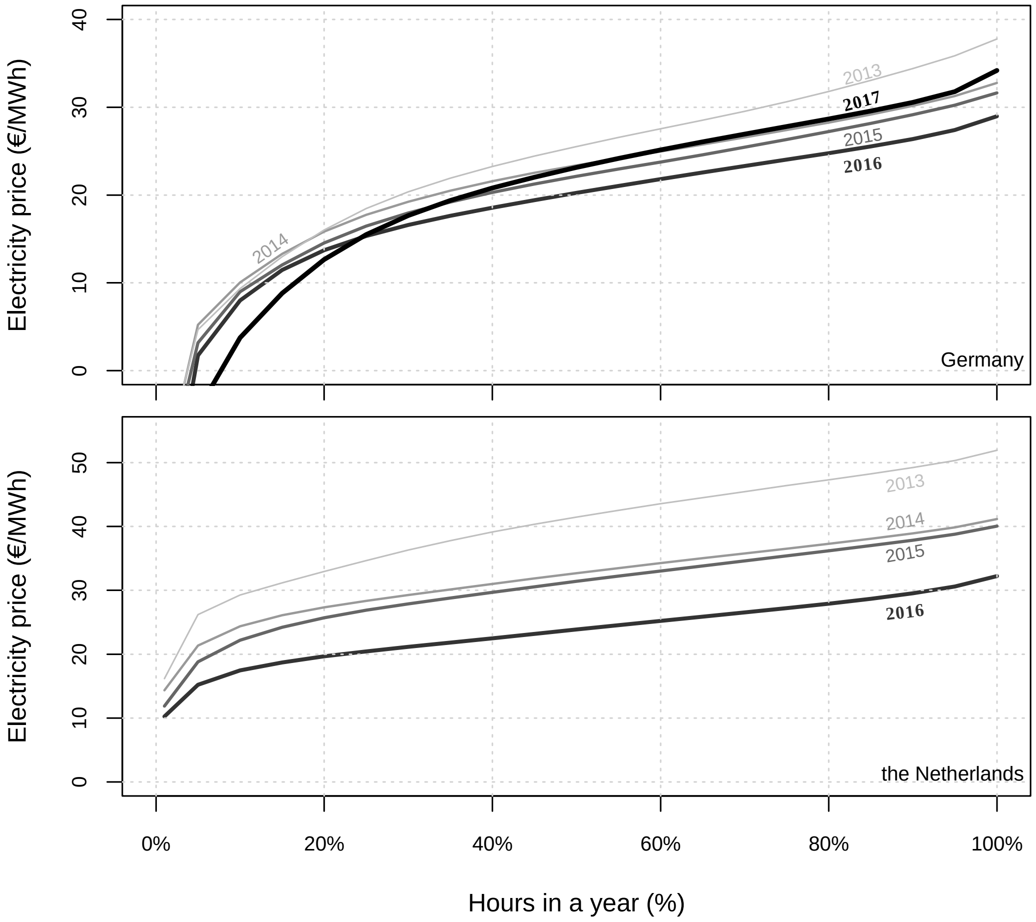

Figure 5: Average electricity prices in cheapest x% of the hours in Germany and the Netherlands (Bloomberg LP, 2017)

Using Figure 5 one can determine the average electricity price when operating a PtG plant during the cheapest x% hours of the year. This is quite different for the different years in the two countries. For example in the Netherlands, operating a PtG plant during the cheapest 20% of the year in 2013 gave an average electricity price of 32.96 €/MWh, which is already higher than the average electricity price over the whole year 2016 (32.24 €/MWh). Operating the plant during the cheapest 20% of the hours in this year gave an average price of only 19.66 €/MWh.

It is interesting to see that whereas Germany shows a significant amount of hours with negative electricity prices, these negative prices do not occur at all in the Netherlands. The share of the negative prices is, however, also in Germany limited. Taking into account the cheapest 5% of the hours, the average electricity price is already positive for the years 2013 – 2016. In 2017, more negative prices occurred: the price was negative during 145 hours (1.7% of the time) and for the cheapest 5% of the hours, the average price was found to be -4.52 €/MWh.

3.3 CO_2

In addition to water and electricity, a PtM plant also requires CO_2 as a feedstock. There are many different sources of CO_2 that can be used for methanation of hydrogen. The costs for capturing CO_2 are not easy to define in general and depend strongly on the concentration of CO_2 in the source stream and the underlying process. It is therefore reasonable to relate the costs of CO_2 to the CO_2 source. The options can be grouped in three main categories: fossil sources, biogenic sources and ambient air. The different sources of CO_2 and their (acceptability as) input for green gas production have been discussed already in deliverable 8.1.[z] Here, the different sources of CO_2 will be shortly discussed with regard to potential and costs. An extensive overview of the availability of CO_2 sources within Europe will be given in a later stage of the STORE&GO project in Deliverables 8.7[aa] and 8.9[bb].

3.3.1 Fossil sources

Fossil sources of CO_2 include power plants and industrial sources. There are on-going discussions whether or not fossil sources of CO_2 should be accepted as an input for green gas production. Some state that the origin of the CO_2 is not important and does not influence the sustainability of the produced gas (e.g. (Meier et al., 2017). The CO_2 is only reused before it is eventually emitted to the atmosphere and therefore it does not matter whether this reused CO_2 is from fossil or biogenic origin, as long as it is not produced specifically for the methanation process. Others claim that total greenhouse gas emissions are lower when biogenic or atmospheric CO_2 sources are used instead of fossil sources (e.g. (Meylan et al., 2017). An argument that is often used against the use of fossil CO_2 sources is that this could legitimize the continuation of a fossil plant. This could be an issue, for example, for coal-fired power plants. For some industrial sources, however, this might be less relevant when they have no green alternatives and will therefore continue to exist and emit CO_2 anyway. An example of this is the cement production industry. Valorisation of these unavoidable CO_2 sources can help decreasing industrial GHG emissions (Meylan et al., 2017).

Most CO_2 sources from industry require a CO_2 capture and upgrading of the captured gas to remove poisoning trace gases before it can be used. This upgrading is expensive and decreases the energy

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[z] Deliverable 8.1: Exploring the future for green gases. Due date 31 August 2017.

[aa] D8.7: Report on the data sources, assumptions and extrapolation methods used for GIS-based modelling of the EU-wide potential for PtG-based storage and “green gas” delivery. Due date: 30 April 2019

[bb] D8.9: Report on a detailed regional analysis of optimal locations of PtG activity, regional drivers and barriers, visualised in a EU-wide set of power-to-gas maps. Due date: 29 February 2020

efficiency. Power plants and industrial sources with low CO_2 contents are therefore not preferred. Some industrial sources have relatively high-purity CO_2 streams, which would be more suitable (Götz et al., 2016).

(Bains et al., 2017) give an overview of capturing CO_2 from industrial sources, including cost estimations for the top CO_2 emitting industries in the United States. Highest costs were calculated for natural gas plants (75 – 100 \$/tCO_2), which is mainly due to the low CO_2 content in the flue gases (3 – 5 mol%). For cement production industries, costs were estimated at 26 – 42 \$/tCO_2 and for iron and steel production at 31 – 35 \$/tCO_2. Lowest costs were estimated for industrial processes with CO_2 purity >95%, which include natural gas processing, ethylene oxide, hydrogen, ammonia and ethanol production. These pure streams basically only need dehydration and compression. Still, capital costs for building a capture system and OPEX costs for operating it are needed, resulting in total costs of 14 \$/tCO_2.

(Reiter and Lindorfer, 2015) evaluate CO_2 sources for PtG applications with regard to capture costs, specific energy requirement and CO_2 penalties. The specific costs for CO_2 capture for industrial processes are further divided into branches of industry. The capture costs for CO_2 from ammonia production range from 23 – 54 €/t CO_2. Capture costs from refineries are higher, about 44 – 94 €/t CO_2. From all the analysed industrial processes, the iron and steel industry have the lowest capture cost at about 16 – 41 €/t CO_2. The cement industry also makes a major contribution to CO_2 emissions. In order to capture this CO_2, costs of about 33 - 69 €/t CO_2 have to be expected. The costs for CO_2 capture form fossil power plants depend on the capture technology (Post-combustion, pre-combustion and oxyfuel) as well on the fuel for the combustion (coal or natural gas).

(NETL, 2014) also present estimates for costs of CO_2 capture from industrial sources. For iron/steel and cement industries with a relatively low purity of CO_2 costs are estimated to be around 100 \$/tCO_2. For industries with high purities of CO_2 (>99%) such as ethanol, ammonia, ethylene and natural gas processing industries, capture costs of CO_2 are estimated to be in the range of 18 – 30 \$/tCO_2.

(Schiebahn et al., 2015) gives some cost estimates for the different kinds of CO_2 based on literature sources and estimates these to be in the range of 20 – 60 €/t for CO_2 from fossil power plants. For CO_2 from industrial processes, no cost estimates are given.

3.3.2 Biogenic sources

Biogenic sources such as anaerobic digesters, bioethanol plants or wastewater treatment plants are very suitable for delivering CO_2 to PtG plants, amongst others, because the sources are widely accepted as an input for green gas production. Biogas upgrading plants with feed-in to the natural gas grid provide a source of otherwise unused CO_2. (Trost et al., 2012) estimate the costs of CO_2 extraction from biogas plants to be roughly 90 €/tCO_2. For this, it is assumed that CO_2 extraction in biogas plants is approximately €0.12 per standard cubic metre of methane for medium scaled plants, following (Fraunhofer Umsicht, 2009), and that the CO_2 fraction in the raw biogas is equal to 40%. Since the effort of removing the CO_2 from the biogas is already done for the retrieval of biomethane, however, costs of the CO_2 for the PtG plant can also assumed to be zero. Using biogas as source for CO_2 it should also be kept in mind that raw biogas CH_4/CO_2-mixture can be used for further methanation without any additional treatment. So if it is not mandatory to separate the feed-in of the biogas from the methanation process, additional costs for CO_2 sequestration can be totally avoided (Trost et al., 2012).

The costs for CO_2 from a bioethanol plant as source behave similarly. Within the fermentation process a high quality stream of CO_2 is accumulated as a by-product. If only considering this as a potential source, the sequestration costs limit to the costs for eventually necessary compression of the gas, which can be assumed to be about 18 €/t CO_2 (Laude et al., 2011). If the bioethanol plant uses cogeneration for energy provision and capturing CO_2 from the cogeneration process should also be considered, costs are between €50 (Möllersten et al., 2003) and €110 (Laude et al., 2011) per ton CO_2 (capturing and compression).

According to (Reiter and Lindorfer, 2015) the costs for CO_2 from biogas treatment and bioethanol production are very low: about 5 – 9 €/tCO_2. This is due to the already high purity, which only requires a drying step for further utilization.

Sewage gas producing wastewater treatment plants as a source for CO_2 have almost the same preconditions as biogas plants. As the incoming sewage gas is also a high quality mixture of methane and carbon dioxide it can also be directly used in a following methanation process without further treatment and therefor no additional costs for sequestration of CO_2. If, out of any reason, the separation of CH_4 and CO_2 is really needed, same costs as mentioned for biogas treatment (90 €/tCO_2) can be assumed.

Limitations on the direct usability of biogenic CO_2 streams from fermentation processes for methanation may result from gas contaminants and trace substances known as catalyst poisons (H_2S, siloxanes, etc.).

(Schiebahn et al., 2015) gives some cost estimates for the different kinds of CO_2 based on literature sources and assume that CO_2 from biomass is available free of charge, except for potential transportation costs.

Two STORE&GO demonstration sites use a biogenic source of CO_2 for their methanation process. The plant in Falkenhagen (Germany) uses CO_2 from a sugar / bio-ethanol factory using material of 100% biogenic origin. The produced CO_2 is usually sold to the beverage industry. The CO_2 is transported over a distance of 300 km by trucks and stored on site in liquid form in a CO_2 storage tank. Costs for the CO_2 are reported to be ~130 €/tCO_2. The sugar / bio-ethanol plant is usually selling the CO_2 to the beverage industry and therefore asks a price. Additionally, the transport over 300 km by trucks also brings costs. The plant in Solothurn (Switzerland) uses CO_2 from a wastewater treatment plant that is transported through a pipeline of 2.55 km to the methanation plant. The CO_2 is received here at no charge, as it would be sent to the incineration if the PtG plant did not use it.

Although biogenic plants seem to be the most ideal sources of CO_2, the average size of the CO_2 sources is rather small. For example, (McKenna et al., 2018) studied possible CO_2 sources for PtG in the Baden-Württemberg region in Germany and found a total of 915 biogas plants in the area with total CO_2 emission of 835 kt CO_2 per year. On average, the plants thus emit less than 1 kt CO_2 per year. To compare: a PtG plant with a 1 MW electrolyser (efficiency 67%) and methanation reactor (efficiency 77.9%) would need 0.8 kt CO_2 per year for full time operation. Thus, an average biogas plant would not be able to deliver CO_2 for plants larger than 1.1 MW. According to the map shown by (McKenna et al., 2018), sizes of biogas and biomethane plants vary between 100 and 12,300 t CO_2 per year. The largest plants are thus able to supply a PtG plant with a 15 MW electrolyser, assuming full time operation and the same efficiency assumptions.

3.3.3 Ambient air

Another possibility to gather CO_2 for a PtG plant is ambient air extraction. The main advantage of this option is that there are no site restrictions and that it can in principal be done everywhere. The concentration of CO_2 in the atmosphere is, however, only roughly 400 ppm and it needs to be concentrated to an almost pure form, which is a very energy intensive and expensive process (Schiebahn et al., 2015).

The technology is still in an early development phase and cost estimates vary widely: e.g. (Trost et al., 2012) estimates the costs at ~160 $/t whereas (House et al., 2011) estimates them at ~1000 $/t. The latter estimates future costs to decrease to ~300 $/t. (Lackner, 2010) estimates future costs of 30 $/t$CO_2$. All these sources refer to capture by sorption processes. Condensation in cryogenic distillation processes or separation from air with membranes are other options, but these options are presumed to be intense in energy usage (Trost et al., 2012).

A report by (Socolow et al., 2011) differentiates between the costs of pure CO_2 separation and the costs of reducing CO_2 in the atmosphere, which means, taking into account the energy sources required for the separation. The reported costs are 610 – 780 \$ and 430 – 550 \$ per ton CO_2, respectively. In an optimization process, these costs could subsequently be reduced by another 10 – 20 % (Mazzotti et al., 2013). According to (Zeman, 2014) the costs reported by (Socolow et al., 2011) can be reduced to approximately 310 \$/tCO_2 by adjusting the separation process.

In the STORE&GO demonstration plant in Troia (Italy) CO_2 is extracted from ambient air. To do so, thermal energy and electricity is required. The company Climeworks who is developing the air extraction plant in Troia expects CO_2 production costs to be in the range of 80 – 120 €/t CO_2 for commercial scale plants with favourable energy costs (low price for waste heat of 100 °C).

The most extreme estimates differ by an order of magnitude. (Ishimoto et al., 2017) also found this huge difference between costs estimates of direct air capture (DAC) reported by academic literature and the companies building the technology. The authors state that none of them are more prone to cost biases but the reason for the big discrepancies is the deep uncertainty in the costs.

3.3.4 Cost comparison

The costs for CO_2 capture from the literature screening for different CO_2 sources (fossil, biogenic and ambient air) are summarized in Figure 6. The green highlighted CO_2 sources in the figure are utilized for capturing CO_2 within the STORE&GO demonstration plants.

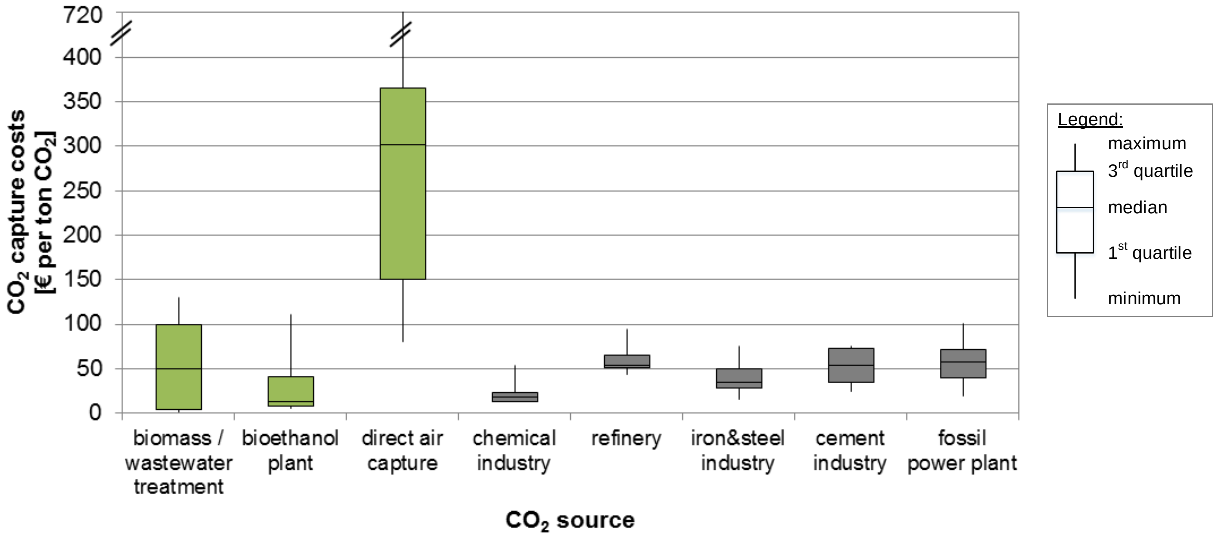

Figure 6: Costs for CO_2 capture from the three STORE&GO demonstration sites and different literature sources, including: (Bains et al., 2017; House et al., 2011; Lackner, 2010; Laude et al., 2011; Mazzotti et al., 2013; NETL, 2014; Reiter and Lindorfer, 2015; Schiebahn et al., 2015; Socolow et al., 2011; Trost et al., 2012; Zeman, 2014)

In general, with the exception of direct air capture, the costs for CO_2 capture are roughly 50 €/t CO_2. However, it must be mentioned that for biomass/wastewater and bioethanol the costs can also be significantly lower, about 5 €/t CO_2. The costs for CO_2 capture from chemical industry can also be lower due to the already high purity of CO_2 in the gas stream. In contrast, the costs for CO_2 from direct air capture are comparatively high and varies in wide range of about 150 – 360 €/t CO_2. The reason for this is the still early stage of development as well as the low concentration of CO_2 in the air, which requires a high amount of energy for the separation. On the other hand, for direct air capture there is still a high potential for development in reducing costs.

4 Conclusions and discussion with regard to the potential of PtG

This report gives an overview of current costs of PtG plants. Combining all current costs, it is possible to calculate the methane production costs of PtG.

Figure 7 shows the methane production costs plotted against the yearly production hours (as a percentage of the year), using the data presented in this report with chemical methanation (which is the cheapest option of the two in terms of investment costs, as presented in Table 12) and electricity prices in Germany in 2016 (the lowest electricity prices as shown in Figure 4 and Figure 5). For CO_2, costs of 50 €/tCO_2 are assumed. The calculations show the overall methane production costs over a project lifetime of 20 years and include the replacement costs of the electrolyser stack (assumed to be 30% of the total electrolyser system costs) after 10 years. A discount rate of 6% was used for the calculations. Very low yearly operating hours (<15% of the year) are excluded from the graph since total methane production costs become very high there (6.78, 13.20, 65.08 €/m^3 for operating the plant during 10, 5 or 1% of the year respectively).

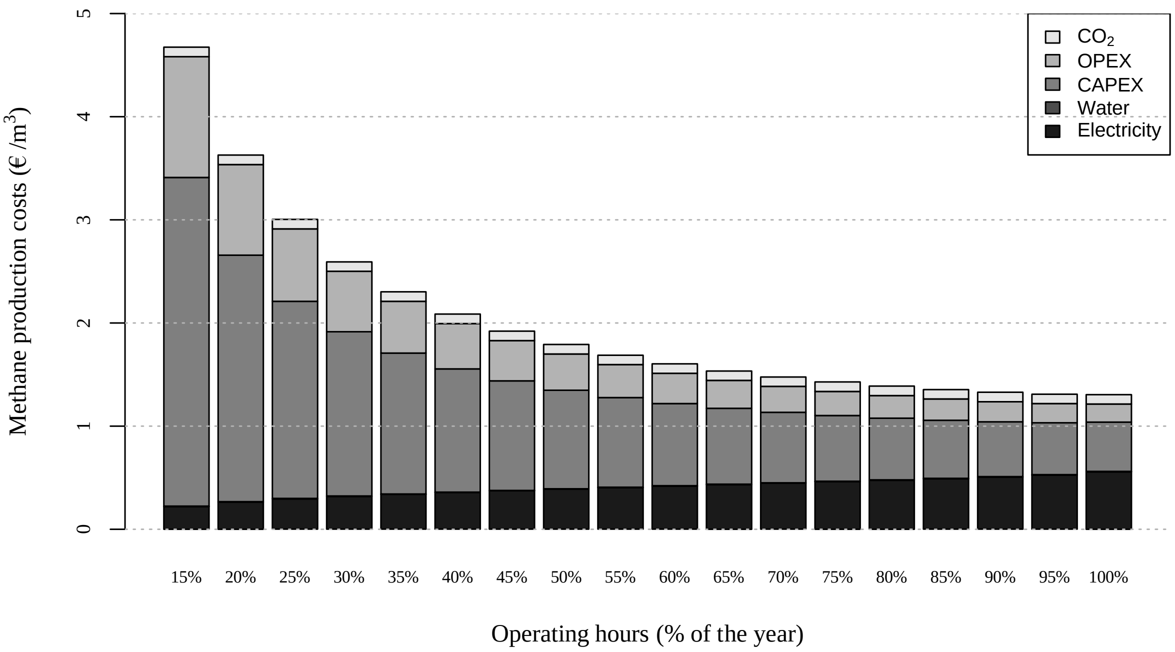

Figure 7: Methane production costs against yearly operating hours (% of the year) using all cost assumptions presented in this report and DA electricity prices of Germany in 2016.

Figure 7 shows that for full time operation of a PtG plant, methane production costs become 1.30 €/m^3. These costs are roughly divided into 43% electricity costs, 37% CAPEX and 13% OPEX and 7% for CO_2. For lower operating hours, the share of CAPEX and OPEX increases, as these fixed yearly costs need to be divided among a smaller amount of yearly methane production. Electricity costs become lower if only cheap electricity can be purchased, but this does not compensate for the increased burden of CAPEX and OPEX on the production costs. Costs for water are negligible in all cases (adding a maximum of 0.3% to the total costs).

The production costs of methane from PtG plants can be compared to current costs of natural gas. In the Netherlands, for example, natural gas costs for households are roughly 0.63 €/m^3. Out of this, 0.25 €/m^3 is paid for the gas, the remaining is taxes (Milieu Centraal, 2017). The costs presented in Figure 7 are thus significantly higher than the current natural gas price, indicating that at the moment, methane from PtG cannot compete with natural gas. Natural gas does not consist entirely out of methane – the Dutch (Groningen) gas contains for example roughly 83% methane (GasTerra, 2018) – but even when this is taken into account methane from PtG cannot compete with fossil methane.

Even when a (significant) price increase is taken into account for the green character of the produced gas (assuming renewable electricity is used and the source of CO_2 is also accepted for green gas production) as opposed to the fossil methane, it cannot compete.

As can be seen in Figure 7, the methane production costs are very dependent on the CAPEX – especially for lower operating hours. Table 12 shows that the electrolyser is by far the most expensive component of a PtG plant. A reduction in the electrolyser investment costs would thus significantly impact the methane production price.

The electrolyser technology is currently still in development. Many literature sources expect significant reductions in costs in the (near) future. Some literature sources – e.g. (Bertuccioli et al., 2014) and (Schiebahn et al., 2015) – expect that the large difference in investment costs between PEM and alkaline electrolysers will disappear and that both technologies might reach investment costs of 500 – 600 €/kW$_{el}$ – half that of today costs of alkaline electrolysers. (Noack et al., 2014) estimates future (~2030) investment costs for 100 MW alkaline and PEM electrolyser systems to be about 520 and 300 €/kW respectively. (Schenuit et al., 2016) estimate future (beyond 2030) costs of both alkaline and PEM electrolysers to be 700 €/kW.

A detailed analysis of potential electrolyser cost reductions will be given in a later Deliverable in the STORE&GO project[cc]. Here, we shortly illustrate the effect of a decrease by half for the electrolyser CAPEX (from 1180 to 590 €/kW) in Figure 8.

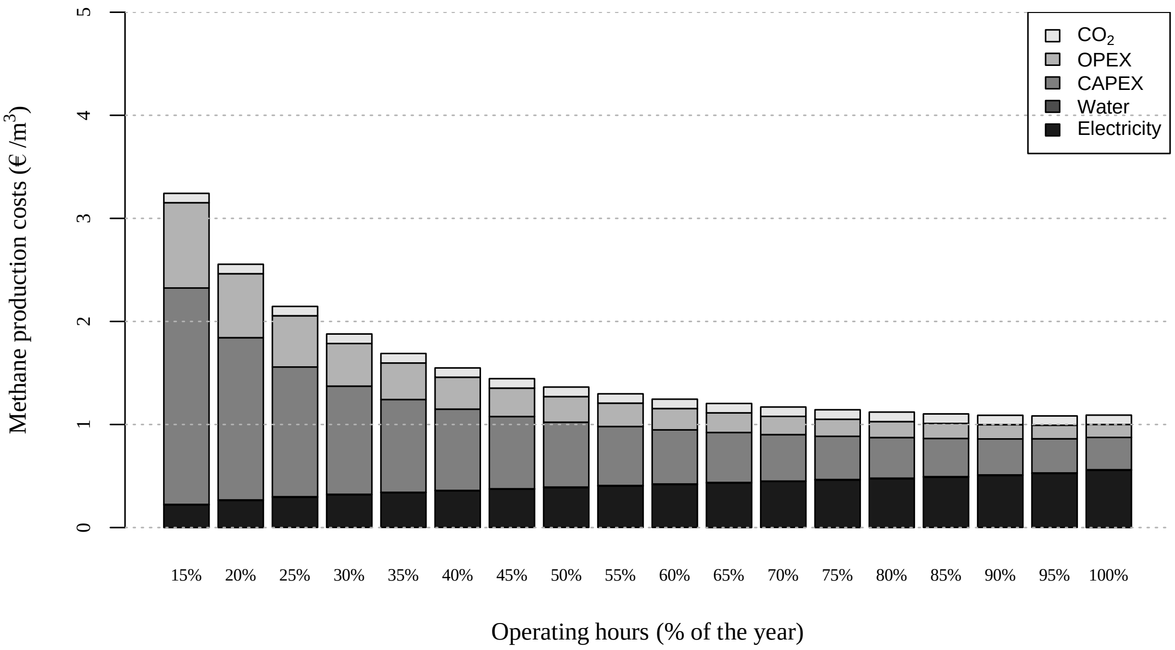

Figure 8: Methane production costs against yearly operating hours (% of the year) similar as in Figure 7 but with reduced electrolyser investment costs (from 1180 to 590 €/kW).

The methane production costs are decreased significantly, especially for low operating hours. For continuous operation of the plant, methane production costs are decreased from 1.30 to 1.09 €/m^3 (a reduction of 16%). Operating the plant for 95% of the time, however, leads to (slightly) lower production costs of 1.08 €/m^3, but this is still more than four times the current price of natural gas. For full time operation, the costs are roughly divided into 51% electricity costs, 29% CAPEX, 11% OPEX and 8% CO_2. Costs for water add only 0.4% to the total costs at most.

[cc] D7.5: Report on experience curves and economies of scale. Due date: 31 October 2018

Summarised, producing methane in a PtG plant cannot compete with natural gas prices today, even when electrolyser investment costs decrease by half. Although a PtG plant could profit from low – or even negative – electricity prices during times of high shares of renewable electricity generation, the high investment costs of the plant require intensive use of the plant to amortize the costs over larger production volumes. It is important to note, however, that a PtG plant will operate only when the marginal costs do not exceed the marginal revenues – assuming the plant does not need to operate continuously because of e.g. a lack of storage capacity or delivery obligations. Thus, when the electricity price becomes too high such that the methane revenues cannot cover it, the plant will not operate even though overall higher production volumes would lower the CAPEX and OPEX per kg produced methane.

In the future, many things could change that would improve the business case of PtG. Besides a reduction in electrolyser investment costs, also the CAPEX of other parts of the PtG plant could decrease. The electricity prices could also change: a further reduction in the average price but also a different pattern with more hours with low prices would improve the PtG business case. One has to note, however, that electricity prices used in the calculations presented here are already the lowest prices that have been observed in recent years (DA prices in Germany 2016 without any taxes). Future electricity prices are difficult to predict and both further decreases and increases are expected in literature. In general, it is expected that electricity prices decrease as the share of renewable energy power generation with very low marginal costs increase. In the long-run, however, the growing share of variable renewables will influence the power generation capacity mix leading to more balanced prices (Green et al., 2011; Helistö et al., 2017).

Another important factor for the PtG business case that is not covered in this report is the revenue that can be received for the produced methane. The methane has to compete with natural gas – although the revenues could be higher due to the green character of the gas, as opposed to the fossil nature or natural gas. The market perspective of green gases is a topic that will be further investigated in deliverable 8.5[dd]. In the future, natural gas prices are expected to increase – both because of an increased commodity price as well as due to higher CO_2 emission prices. Methane produced in a PtG plant can become more competitive with natural gas under these circumstances.

[dd] D8.5: Study describing the short, medium and long-term perspectives of various market segments for “green gases”. Due date: 31 October 2018

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