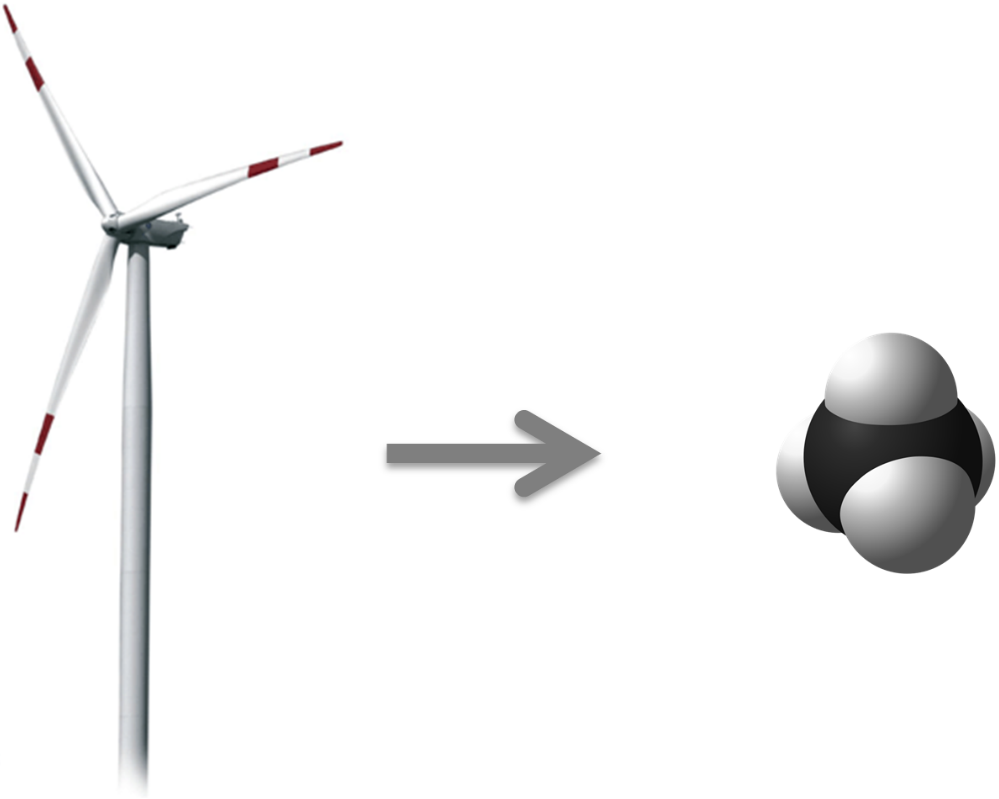

# Integration of Power-to-Gas in Gasendal and GoBiGas

*Master's Thesis within the Sustainable Energy Systems programme*

JONATAN AGERSBORG, EMIL LINGEHED

Department of Energy and Environment
*Division of Energy Technology*
CHALMERS UNIVERSITY OF TECHNOLOGY
Göteborg, Sweden 2013
Report No. T2013-396

MASTER’S THESIS

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## ABSTRACT

Power-to-Gas, which is the term used to describe the use of electrolysis and Sabatier reactor systems to react $H_2$ and $CO_2$ and form $CH_4$, is a technology that has received a great deal of attention recently. The produced $CH_4$ can be used in a variety of applications, including fuel for vehicles and energy storage in existing natural gas networks. Research is currently being conducted on the performance of this technology, and several demonstration plants have been built. However, few evaluations have been made on the implementation of power-to-gas in existing biogas production plants. This thesis will look at the technical and economic aspects involved in integrating the power-to-gas technology in a biomass gasification plant (GoBiGas) and a biogas upgrading plant (Gasendal).

The technical evaluation is carried out using the process engineering software Aspen Plus. The Sabatier reactor system is modelled and optimized with regard to material and reactant composition restrictions, and Pinch analysis is carried out to evaluate the potential for heat recovery. An economic evaluation is made to check the profitability of the implementation with respect to current energy market conditions, and a sensitivity analysis is done to conclude what future market changes and technology conditions may affect the economic outcome.

In conclusion, the implementation of power-to-gas is profitable for neither of the two applications with current market and technology conditions. However, investment costs are predicted to fall and SNG prices are likely to rise in the future, which would benefit the economic outcome of the implementation.

*Keywords*: Power-to-gas, Sabatier Reactor System, Aspen Plus, GoBiGas, Gasendal

Integration av Power to gas i Gasendal och GoBiGas
Examensarbete inom masterprogrammet *Sustainable Energy Systems*
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Chalmers tekniska högskola

## SAMMANFATTNING

”Power-to-Gas”, namnet på en process i vilken el omvandlas till metangas, har fått en hel del uppmärksamhet på senare tid. I processen matas el in i en elektrolyser som genererar vätgas; vätgasen och koldioxid förs in i ett Sabatier-reaktorsystem som producerar metangas. Den producerade metangasen kan användas på olika sätt, framför allt som ett alternativt fordonsbränsle men även för att lagra el-energi i gasnätet. Forskning för att utvärdera den här teknologin sker för tillfället, flera demonstrationsanläggningar har byggts. Det har dock inte gjorts studier på integration av ”power-to-gas” processen i existerande biogasanläggningar. Det här arbetet syftar till att undersöka de tekniska och ekonomiska aspekterna relaterade till att integrera ”power-to-gas” processen i en förgasningsanläggning (GoBiGas) och i en biogasuppgraderingsanläggning (Gasendal).

Den tekniska analysen utförs i processflödesprogrammet Aspen Plus. Sabatier-reaktorn modelleras och optimeras med hänsyn taget till material- och gaskompositionsbegränsningar. En Pinchanalys genomförs för att undersöka potentialen för värmeåtervinning. En ekonomisk analys utförs för att undersöka lönsamheten för processen med avseende på dagens prisnivåer. En känslighetsanalys görs för att undersöka vilka framtida prisförändringar som skulle krävas för att göra processen lönsam.

Sammanfattningsvis, att integrera “power-to-gas” processen är inte lönsamt för något av de två alternativen med nuvarande marknad och teknologi. Att integrera processen hjälper ekonomin men det är fortfarande långt från tillräckligt för att nå lönsamhet i processen. I en framtid där värdet på den producerade metangasen stiger och investerings kostnaderna sjunker skulle processen kunna vara lönsam.

*Nyckelord*: Power-to-gas, Sabatier reactor system, Aspen Plus, GoBiGas, Gasendal

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Göteborg, June 2013

Jonatan Agersborg & Emil Lingehed

# Contents

# Notations

## Letters

| | |
|---|---|
| $P$ | Pressure of the system, bar |
| $P^0$ | Ambient pressure, bar |
| $X_i$ | Molar fraction of component i |
| $R_u$ | Universal gas constant, J/mol K |
| $T$ | Temperature, K |
| $\Delta G_T^0$ | Standard state Gibbs function change at temperature T, J/mol |

## Abbreviations

| | |
|---|---|
| *Bio-gas* | Methane produced from biological sources |
| *E-gas* | Methane produced by a power-to-gas process |
| *RME* | Bio-diesel |
| *PEM* | Proton Exchange Membrane |
| *SNG* | Synthetic natural gas, including both bio-gas and e-gas |

# 1 INTRODUCTION

A number of challenges and technical aspects need to be overcome to allow our future energy system to be environmentally and economically sustainable. Certain challenges, such as the shift from fossil to renewable energy sources and the reduction of anthropogenic $CO_2$ emissions, are already being addressed. However, the increasing share of renewable energy from intermittent sources is introducing new challenges in the energy system.

Solar and wind power have seen a large expansion the last decade, while the existing electricity network remains largely unchanged. This has left the current electrical energy system un-optimized and in need of adjustments. The electricity transmission system needs to be adapted to smaller local energy production sites instead of the larger scale production sites used today. Energy storage solutions need to be implemented or energy consumption patterns need to be adapted to compensate for the fluctuations in intermittent energy production.

Several technologies have been developed and implemented that can replace fossil fuelled electricity production plants, but the situation is more difficult in the transportation sector. Liquid fossil fuels dominate the source of energy in the transportation sector due to their relatively easy transport capabilities and high energy density. The transportation sector therefor possesses a large potential for finding innovative energy solutions, and renewable methane is a promising alternative.

## 1.1 POWER-TO-GAS

The concept of power-to-gas is the transformation of electrical energy to SNG (synthetic natural gas) using an electrolyser and a Sabatier reaction. The electrical energy can essentially come from any source, the ideal situation being if the electrical energy comes from intermittent energy sources at times when production is high and consumption is low, as is the case sometimes in energy systems with large amounts of wind and solar power production. The electricity is used to split water in an electrolyser, and the resulting $H_2$ is used to produce $CH_4$ in a Sabatier reactor using some source of $CO_2$. The produced $CH_4$, also called e-gas, can be injected and stored in existing natural gas grids or used as a fuel for transportation. The concept is shown in Figure 1.

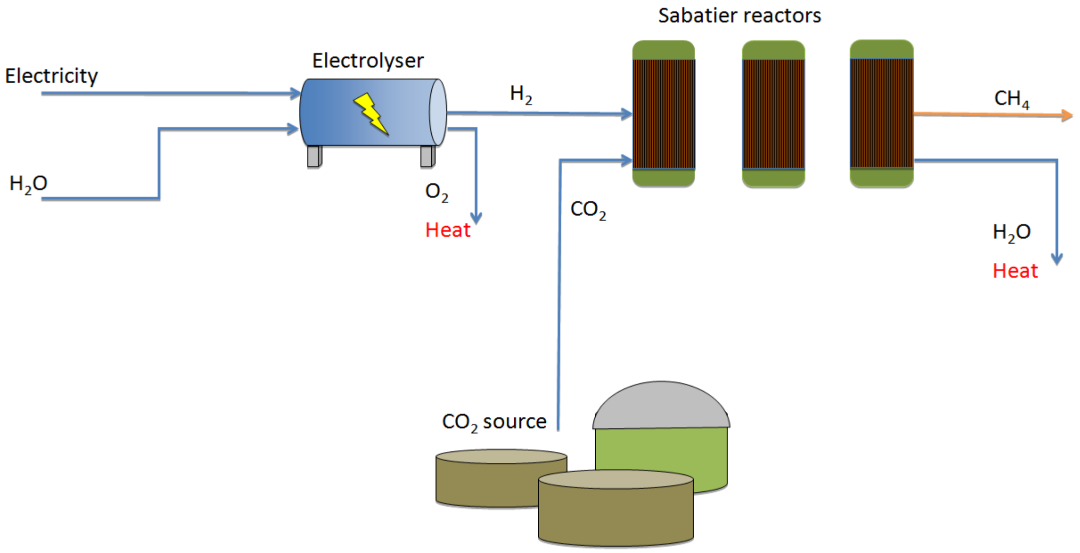


**Figure 1.** An overview of the power-to-gas process.

The implementation of power-to-gas is currently being investigated in Germany and Denmark, with several research facilities investing in pilot plants of various scales. Audi has built a 6,3 $MW_{el}$ production facility using this technology to produce sustainable fuel for a fleet of gas-powered cars, and SolarFuel has built several other demonstration plants (SolarFuel GmbH 2013). The technologies involved in the power-to-gas process are relatively well tested units. Electrolysers exist on the current market and Sabatier reactors have been developed for space missions and are very similar to the existing methanation reactors involved in gasification processes (Haldor Topsøe n.d.).

## 1.2 PROBLEM STATEMENT

This thesis analyses the technical aspects of e-gas production using a power-to-gas process, and evaluates the feasibility of integrating the process in two existing bio-gas production plants. The two plants investigated are the GoBiGas plant, a biomass gasification plant, and the Gasendal plant, a biogas upgrading plant. The technical aspects are adjusted to optimize e-gas production using heat recovery and process modelling techniques. The goal is to determine if it is economically feasible to produce e-gas given present energy market characteristics. The process is then analysed in an alternative energy system scenario with a higher penetration of intermittent energy.

## 1.3 SCOPE

The purpose of the thesis is not to evaluate the performance of individual equipment involved in the power-to-gas process, but instead to evaluate the integration of the system as a whole with the existing production plants.

The thesis does not consider the distribution or storage of the produced methane, as the required infrastructure is already available. Also, the aim of the thesis is not to derive new energy system models where the methanation process would be economically feasible but instead to evaluate the process with respect to existing electricity price curves for 2011 and 2012, and also to evaluate the process with respect to an existing electricity price prediction model for 2022.

## 1.4 GASENDAL

The Sewage treatment plant in the Gothenburg area, Gryaab, produces bio-gas by anaerobic digestion of sludge. The bio-gas produced contains about two thirds methane. In order to inject the bio-gas to the natural gas grid it has to be upgraded (Jacobsson 2010). This upgrading of the bio-gas is what is being done at the Gasendal plant.

The upgrading process contains several steps including $H_2S$ removal, $CO_2$ removal, compression, drying and propane addition. An overview of the process can be seen in Figure 2. The plant has a maximum capacity of 1600 $Nm^3$ bio-gas/hr with an average flow of 1018 $Nm^3$/hr (Eklund 2013).

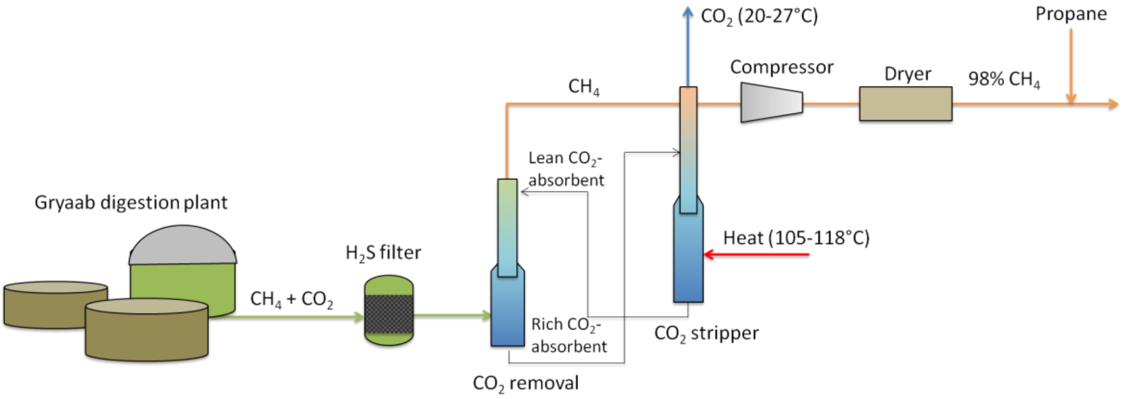


**Figure 2.** Gasendal bio-gas upgrading process (Purac n.d.), (Eklund 2013).

The first step of the process is the $H_2S$ removal, which is done with an activated carbon filter. Due to very low concentrations of $H_2S$ in the bio-gas from Gryaab, changing the filter is rarely needed (Zinn 2013). The concentration of $H_2S$ after the filter at Gasendal is lower than 5 ppm (Karlsson 2013). The catalyst used in the power-to-gas process is highly sensitive to $H_2S$ and therefore it is important to monitor the filter's condition carefully if the power-to-gas process is to be integrated at Gasendal.

The following $CO_2$ removal is done in a counter current packed absorption column. The bio-gas enters the column at the bottom and $CO_2$ absorption liquid is sprayed from the top. The liquid absorbs $CO_2$ (and $H_2S$ if any has slipped through the filter) and is then pumped to the $CO_2$ stripper. Here, the liquid is heated above its boiling point using steam produced from burning natural gas and the $CO_2$ is released from the liquid. The $CO_2$ is cooled before it is released to the atmosphere to recover the $CO_2$ absorption liquid (Karlsson 2013).

The $CO_2$ is then released in the atmosphere and the $CH_4$ is compressed to a pressure of 5 bar(a). The compression is done in two Mehrer compressors TEW90-Ex and they together have a capacity of 1064 $Nm^3$/hr, determining the maximum $CH_4$ flow of the plant (Strandberg 2013).

After the compression, the $CH_4$ is dried and contains about 98% $CH_4$. Propane is then added to reach the required lower heating value from the grid specification (Purac n.d.). Table 1 shows a summary of the key values from the Gasendal plant.

**Table 1.** Summary of key operation values from Gasendal.

| | **Unit:** | **Value:** |
|---|---|---|
| **Bio-gas flow to Gasendal** [a] | $Nm^3$/hr | 1020 |
| **Methane flow** [a] | $Nm^3$/hr | 680 |
| **Carbon dioxide flow** [a] | $Nm^3$/hr | 340 |
| **Heat required in $CO_2$ removal** [b] | $kJ/Nm^3_{CO2}$ | 7450 |
| **Heat required in $CO_2$ removal, on average** | kW | 700 |
| **Temperature of required heat** | °C | 110 |
| **$CO_2$ temperature** | °C | 23 |
| **Methane production, on average** | MW | 6,4 |
| **Annual methane production** | GWh/year | 60 |

[a] *average flow,* [b] *based on average flows, see Appendix A for more details*

## 1.5 GoBiGas

A 20 $MW_{bio\text{-}gas}$ biomass gasification plant is currently under construction in the Gothenburg area (July 2013). The plant utilizes indirect steam gasification to transform forest residues into methane using a number of different technologies that will be explained in greater detail below.

Allothermal (or Indirect) gasification is a technology where biomass is thermally gasified in a reactor using steam and heat from a separate furnace. The primary product of this process is syngas that consists of mainly $H_2$ (40%), CO (20%), $CO_2$ (20%) and $CH_4$ (10%). The syngas is then processed in a methanation stage where the $H_2$ and CO is converted into $CH_4$. Some $CO_2$ will remain when full conversion to $CH_4$ is achieved and this $CO_2$ is usually separated and reused within the process and later emitted to the atmosphere. Figure 3 shows a schematic view of the overall process and divides the process into the gasification process, the top half, and the methanation process, the bottom half.

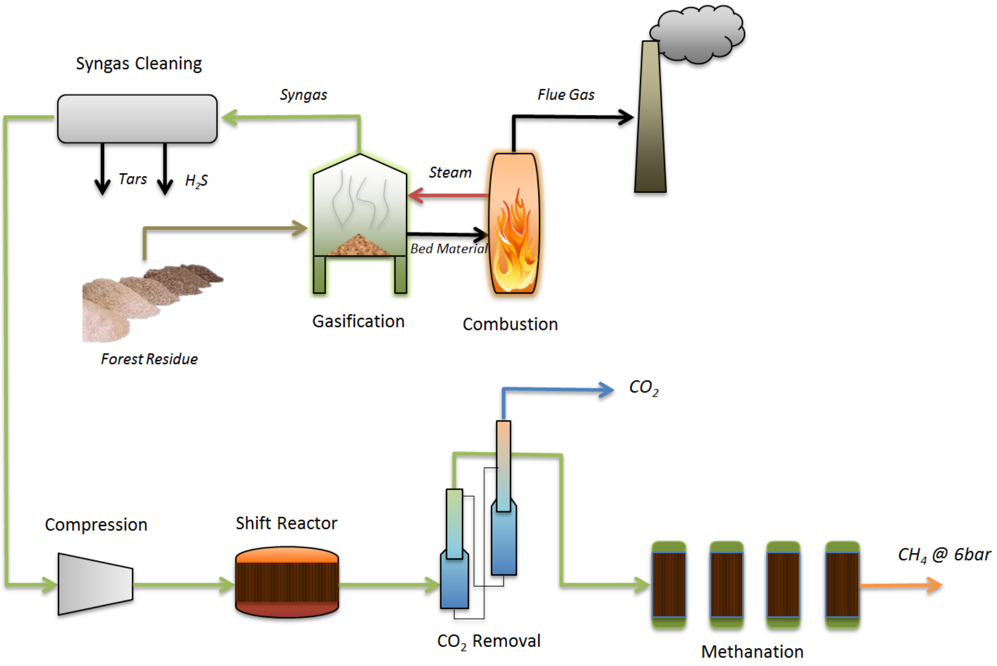


**Figure 3.** Schematic of the GoBiGas process.

The 20 $MW_{SNG}$ plant uses 32 MW of wood residues, 3 MW electricity and 0,5 MW RME during normal production and produces, apart from the 20 $MW_{bio\text{-}gas}$, an excess of heat, where 5 MW is at a high enough temperature to be directly heat exchanged to the district heating network and 6 MW is upgraded at the heat pump plant prior to being distributed as district heating. The produced bio-gas consists of 95% $CH_4$ and is distributed via the local natural gas grid.

# 2 THEORETICAL BACKGROUND

## 2.1 ELECTROLYSER

An electrolyser is a device that splits water into hydrogen and oxygen by using electricity, shown in equation (1). The basic design of an electrolyser consists of two electrodes (cathode and anode) and an electrolyte (Larminie and Dicks 2003).

$$H_2O + 2e^- \rightarrow H_2 + ½O_2 \tag{1}$$

Ideally 39 kWh of electricity and 8.9 kg of water are required to produce 1 kg of hydrogen at 25°C and 1 atm pressure (Harrison and Levene 2008). The enthalpy change when water is formed is 286 kJ/mol$_{H2}$ at 25°C and 1 atm pressure (Nave 2012). In a real unit, there are losses and the efficiency of an electrolyser system ranges between 50 and 80% (Harrison and Levene 2008).

Electrolysers are categorized by the type of electrolyte they use. There are two types of electrolysers that are commercially available today Alkaline and Proton Exchange Membrane (PEM). The main difference between the two types is that the alkaline electrolysers' electrolyte is a liquid and that the PEMs' electrolyte is a solid.

Alkaline electrolysers are the most common type in large applications where a large amount of hydrogen is produced. They generally have lower investment cost than PEM electrolysers (Harrison and Levene 2008). The PEM type is faster at adjusting its power consumption than the alkaline type and it also has benefits if the electrolyser is pressurized.

## 2.2 SABATIER REACTION

The Sabatier reaction is an exothermic reaction where $H_2$ and $CO_2$ reacts to form $CH_4$ and $H_2O$ and follows the reaction process shown in equation (2).

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O \qquad \Delta H = -165\frac{kJ}{mol} \tag{2}$$

The exact reaction mechanism is under discussion but the most widely accepted mechanism is the combination of a reversed endothermic water-gas-shift-reaction and an exothermic CO methanation, both which can be seen in equation (3) and (4) respectively (Sterner).

$$H_2 + CO_2 \rightarrow CO + H_2O \qquad \Delta H = 41.5\frac{kJ}{mol} \tag{3}$$

$$3H_2 + CO \rightarrow CH_4 + H_2O \qquad \Delta H = -206.4\frac{kJ}{mol} \tag{4}$$

The overall reaction (2) is favoured at lower temperatures, but due to kinetic limitations, a catalyst needs to be utilized (Brooks, et al. 2007).The implementation of this reaction therefore requires a careful heat management to maintain the reaction at relatively low temperature for a favourable equilibrium composition but at a sufficiently high temperature to overcome activation energies. The equilibrium of the Sabatier reaction is described in greater detail in the next section.

In commercial applications, a nickel catalyst is usually used, due to its selectivity, activity and its price, but other metals such as Rubidium, Rhodium, Platinum, Iron, and Cobalt may also be used. (Kopyscinski, Schildhauer and Biollaz 2010), (Grond 2013), (Hoekman, et al. 2010)

## 2.3 CHEMICAL EQUILIBRIUM OF THE SABATIER REACTION

Chemical equilibrium assumes that the reaction has infinite time to react to reach the equilibrium; it does not take any kinetics into account. One way to determine the equilibrium is to use Gibbs free energy where $K_p$, the equilibrium constant, is an important variable. The equations below (5-6) are the equations related to $K_p$ for the Sabatier reaction (Turns 2011):

$$K_p = \frac{X_{H_2O}^2 * X_{CH_4}}{X_{CO_2} * X_{H_2}^4} \left(\frac{P}{P^0}\right)^{-2} \tag{5}$$

$$K_p = \exp\left(-\frac{\Delta G_T^0}{R_u T}\right) \tag{6}$$

The $K_p$:s at different temperatures for the reaction are calculated with equation (6) and they are presented in Table 2.

**Table 2.** $K_p$ of the Sabatier reaction for the temperature range 127-727 °C (Turns 2011), (M.W. Chase 1986).

| **Temperature** | | **$K_p$** |
|---|---|---|
| **[K]** | [C] | [ - ] |
| **400** | 127 | $2{,}70 \cdot 10^{12}$ |
| **600** | 327 | $7{,}15 \cdot 10^{4}$ |
| **800** | 527 | 7,70 |
| **1000** | 727 | 0,03 |

Knowledge of the $K_p$ of the reaction and equation (5) gives an understanding of how different parameters like composition, temperature and pressure affects the reaction.

### 2.3.1 COMPOSITION

According to Le Chatelier's principle "If a system at chemical equilibrium changes pressure temperature or concentrations, then the equilibrium will shift to the side that would reduce the change" (Nationalencyklopedin 2013). In the Sabatier reaction, when more products (water or $CH_4$) are added to the mixture the conversion of $H_2$ will be disfavoured. Higher concentrations of $CO_2$ would on the other hand increase the conversion of $H_2$. Equation (5) shows that the effect of adding more $H_2$ will be higher than adding $CO_2$ because the $H_2$-fraction contributes with the power of four. Adding more water worsen the conversion of $H_2$ more than adding $CH_4$ because of the squared $H_2O$ factor in the equation.

### 2.3.2 TEMPERATURE

As can be seen in Table 2, $K_p$ is highly dependent on the temperature. Equation (5) shows the effect a change of $K_p$ has on the composition of the system. Since $K_p$ increases with a decrease in temperature, a lower temperature favours $H_2$ conversion. The relation between $H_2$ conversion and the temperature can be seen in Figure 4.

In an adiabatic reactor, the products absorb heat from the reaction which affects the temperature of the reactor. Therefore, the temperature increases also depends on the heat capacities of the products. The heat capacities of the products are shown in Table 3.

**Table 3.** $C_p$ for the four different main species in the Sabatier reaction (Turns 2011).

| Temp [K] | Temp [°C] | $C_{p,\ H2O}$ [kJ/kmol K] | $C_{p,\ CH4}$ [kJ/kmol K] | $C_{p,\ CO2}$ [kJ/kmol K] | $C_{p,\ H2}$ [kJ/kmol K] |
|---|---|---|---|---|---|
| **400** | 127 | 34 | 40 | 41 | 29 |
| **600** | 327 | 36 | 52 | 47 | 29 |
| **800** | 527 | 39 | 63 | 52 | 30 |

### 2.3.3 PRESSURE

The stoichiometry of the reaction shows that the number of moles of the reactants is higher than the number of moles of the products. Therefore, increasing the pressure will shift the equilibrium towards the products. This can also be seen in equation (5) where the $\left(\frac{P}{P^0}\right)^{-2}$ factor indicates a pressure dependence on the final composition. A high pressure has a large effect on the conversion of $H_2$, however the effect will be higher at higher temperatures where the $K_p$ is low. At lower temperatures where the $K_p$ is high, the effect of the pressure will not have as large relative impact, this can be seen clearly in Figure 4.

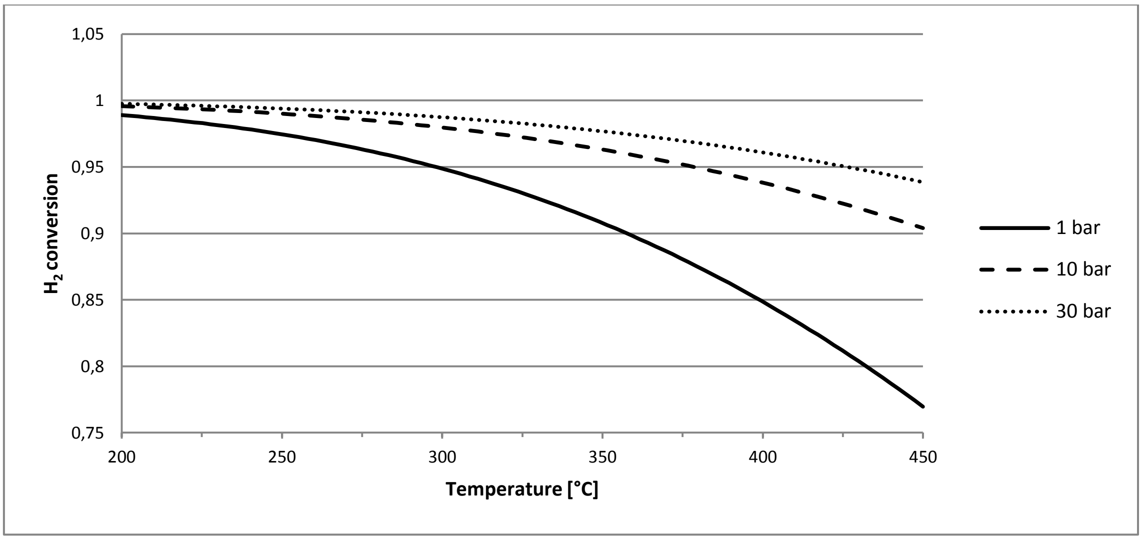


**Figure 4.** Thermodynamic equilibrium of the Sabatier reaction at different temperatures and pressures.

## 2.4 FUNDAMENTALS OF SABATIER REACTOR DESIGN

The three major types of reactors that have been considered for implementing in the power-to-gas process are micro channel reactors, fluidized bed reactors and fixed bed reactors. Micro channel reactors are designed with a large number of very narrow channels with outstanding heat and mass transfer capabilities (Brooks, et al. 2007). These reactors are usually used in applications with a small reactant flow, and are currently only existing on a lab scale and are developed for space applications. Fluidized bed reactors possess good heat and mass transfer characteristics due to the continuous mixing of the catalyst and the reactants within the reactor, and have been implemented for methanation reactors in a few cases. The most widely used reactor design for methanation processes is the fixed bed reactor where the active metal is coated onto the surface of a support. Due to the high rate of heat release, internal cooling is not possible. The reactor system is therefor designed to be adiabatic with cooling after the reactor.

### 2.4.1 CATALYST

The most sensitive element of the reactor is the catalyst and the most widely used catalyst within previous commercial implementations and laboratory tests is the Haldor Topsøe – PK-7R nickel based catalyst. (Hoekman, et al. 2010) The PK-7R is capable of operating at a temperature interval of 190-450°C, but a high temperature catalyst, denoted MCR, is available for applications up to 700°C (Haldor Topsøe n.d.). The advantage of using a high temperature catalyst is that there is a reduced need for temperature control in the reactor and that heat is produced at a higher temperature. However, a higher outlet temperature reduced conversion efficiency and increases demands on the materials used for the process. Carbon formation needs to be avoided and $H_2S$ needs to be removed as these may degrade the nickel catalyst through fouling and poisoning (Bartholomew 2001). Also, the nickel catalyst is sensitive to $O_2$, which practically means that all periods of production stoppage need to be carried out in an $N_2$ atmosphere. (Culmsee 2013).

## 2.5 DESIGN OF SABATIER REACTOR SYSTEM

Two existing designs are used as a starting point for the further analysis of the Sabatier reactor system design, one described by DNV KEMA, and one published by Haldor Topsøe. DNV KEMA is a private research and development company based in Holland currently testing the performance of Sabatier reactors and evaluating the economic potential. Haldor Topsøe is a Danish chemical company with a broad product portfolio including methanation solutions for coal and biomass gasification.

Both companies implement a fixed bed reactor system where the reactors are considered adiabatic due to the lack of internal cooling (Haldor Topsøe n.d.) (Grond 2013). In an adiabatic Sabatier reactor the temperature will increase and it is therefore important to design the reactor system in such a way that the temperature increase is kept under control. A reactor system usually consists of more than one fixed bed reactor and has recycling and water condensation steps to control the temperature and achieve a high conversion of the $H_2$ and $CO_2$ to $CH_4$ (Grond 2013). The reactor system can be built with three separate fixed bed reactors with water condensation after each step, as shown in Figure 5.

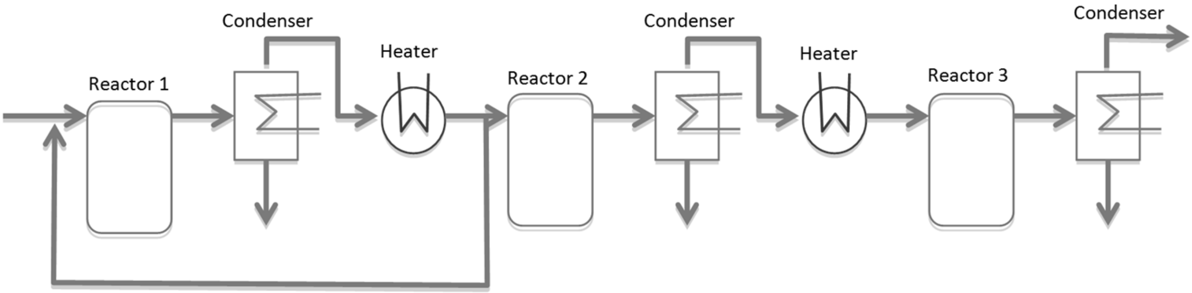


**Figure 5.** An example of a design of a Sabatier reactor system.

The first reactor step in the design in Figure 5 has a recycling to lower the temperature in the first reactor. By implementing a recycling, the concentrations of the reactants in the reactor will be lower, which leads to a lower temperature increase. The water condensation between each step is conducted to cool the product gas between the reactors but also to prevent the reversed reaction from occurring, where $H_2O$ and $CH_4$ are converted back to $CO_2$ and $H_2$, which is recommended by DNV KEMA (Grond 2013). It is also beneficial to remove some water to lower the volume flow of the reactants. On the other hand, some water needs to remain in the reactant flow to avoid carbon formation during the reaction. The optimal water content is investigated in the Aspen simulation and is presented in the section with optimizations of the design parameters.

## 2.6 SHIFT REACTOR

A shift reactor is required to adjust the $H_2$ content of the reactant gases at GoBiGas prior to methanation. The shift reactor performs a water gas shift reaction by adding steam and subsequently decreasing the CO content as described by equation (7). An iron based catalyst is commonly used for the reaction.

$$H_2O + CO \rightarrow H_2 + CO_2 \tag{7}$$

The purpose of the shift reactor is to achieve a ratio of CO to $H_2$ of 1:3 prior to the methanation reactors due to the stoichiometry of the methanation reaction. This is done by bypassing a certain amount of relatively CO-rich syngas around the shift reactor while allowing the remaining stream to flow through the shift reactor.

## 2.7 ELECTRICITY MARKET

In northern Europe, electricity is traded on three markets: the day a head market (called Elspot), the intraday market (called Elbas), and the regulating market. The major difference between them is when the trading occurs. On the Elspot market the trading occurs the day before the actual delivery, on the Elbas market trade occurs until one hour before the actual delivery and on the regulating market it is occurring during the actual delivery. Most of the trading is done on the Elspot market today. Elbas is a market established to put the system into balance before the actual delivery if something unpredicted occurs, ex more wind power production than predicted. The seller can then sell the excess power on this market to avoid having to pay for regulating power for producing more electricity than agreed. (NordPoolSpot n.d.)

The electricity price on Elspot is set by supply and demand for each hour of the day. The sellers and buyers give orders of how much they expect to produce and consume the next day and to what price they are willing to sell and buy the electricity (NordPoolSpot n.d.). The sale price is determined by the variable cost of producing the electricity. Renewable sources like hydropower and wind power has very low variable costs. When the demand is low or when there is a large production of electricity at a low variable cost the system price will be low. If the demand is higher than the production with low variable costs, other production plants will go in to meet the higher demand. These plants could be coal condensing power or gas turbines; they have higher variable costs and will therefore increase the electricity price.

# 3 METHOD

Initially, the main focus is allocated to gathering technical and practical information on the technologies and modelling tools used throughout the technical analysis of the project through literature studies, study visits and personal contact with a variety of sources. When a fundamental theoretical base has been obtained, the process models are created and optimized in Aspen Plus and a thorough analysis of the process integration is carried out using sensitivity and pinch analysis. Finally, an economic analysis is made to determine the profitability of the process and a sensitivity analysis is carried out to identify the scenario that would allow profitability.

## 3.1 PROCESS MODELLING

For the technological analysis presented below, the process modelling software Aspen Plus is used. Aspen Plus can be used to design, test and optimize chemical processes and is in this thesis used to model Gasendal, GoBiGas and the power-to-gas process. It is also used to test the chemical processes with respect to reaction performance, heat recovery optimization, and to determine overall integration potential. In the models of this thesis the Peng-Robinson equation of state with Boston-Mathias modifications are applied.

### 3.1.1 CATALYST

In the Gasendal case, where the Sabatier reactor system is a separated system and the reactors can be designed without existing restrictions. The catalysts are assumed to be the lower temperature, PK-7R type provided by Haldor Topsøe, since the lower temperature allows a higher conversion at low pressures. In the GoBiGas case, the catalyst in the 1st methanation reactor is assumed to be the high temperature, MCR model while the catalyst in the 2nd-4th methanation reactors are assumed to be the PK-7R model, shown in Figure 16. This is in accordance with the outlet temperatures of the design of the methanation process of GoBiGas phase 1, provided by Göteborg Energi.

### 3.1.2 CARBON FORMATION

Carbon formation needs to be minimized to avoid catalyst fouling and a subsequent decrease in the $CH_4$ production. The analysis avoids carbon formation in the Aspen model by adding $H_2O$ and monitoring the formation of solid carbon using equilibrium calculations.

### 3.1.3 REACTORS

The reactors modelled in this analysis are assumed to be adiabatic fixed bed reactors, because they are the most commonly used today for this type of reaction.

The reactor type used in Aspen Plus for this thesis is the RGibbs reactor, a type of equilibrium reactor that calculates the product composition at which the Gibbs free energy of the products is at a minimum. This model is used in cases where the exact reactions are not known but the reactants and possible products can be specified (AspenTech 2003). In the reactor chemical equilibrium is reached, which means the reaction kinetics is not taken into account and the equilibrium is independent of time. For methanation and shift reactions using a catalyst, it is assumed that this simplification is valid for this type of analysis (Seemann 2013).

#### *METHANATION REACTOR*

The methanation reactors are modelled with adiabatic Gibbs reactors. The products of the reaction are specified and the formation of coal is monitored. The reactors calculate the equilibrium of the products and reactants based on the input parameters such as temperature, pressure and composition and the product gases absorb the produced heat of the exothermic reaction. As a result, the temperature of the product gases increase, which leads to a cooling requirement between the reactors.

#### *SHIFT REACTOR*

The modelling of the shift reactor is done with an adiabatic Gibbs reactor where the $CH_4$ is considered inert and therefore does not take part in the reaction. This assumption is done to prevent steam reformation in the modelled reactor. The assumption is considered valid since the shift reactor utilizes a different catalyst than the methanation reactor and therefore no methanation or steam reformation should occur in the reactor. The modelled shift reactor is taken from the *original model* (section 3.1.7), which is based on studies done by Hamelinck (Arvidsson and Heyne 2013).

### 3.1.4 ELECTROLYSER

The electrolyser is modelled with a calculator block in Aspen and is based on data of an atmospheric alkaline-electrolyser from the manufacturer NEL-hydrogen, see Appendix A - Input Data. The model accounts for the electricity consumption, heat generation, $H_2$ production, $O_2$ production and the water consumption.

### 3.1.5 AUXILIARY EQUIPMENT

Both the $CO_2$ and $H_2S$ removal equipment are modelled as separation units where the separation fractions are calculated in a separate Aspen file using a standard amine absorber unit (Arvidsson and Heyne 2013). The drying equipment prior to connection to the natural gas grid is also modelled as a separation unit with an estimated $H_2O$ separation of 98%.

### 3.1.6 PRESSURE DROP

The pressure drop of individual units is estimated rather roughly both from information directly from Göteborg Energi as well as experience from literature review and from similar models.

### 3.1.7 ORIGINAL GOBIGAS MODEL

The modelling of the methanation stage of GoBiGas originates from a model created by Maria Arvidsson and Stefan Heyne, which will be referred to as the *original model* (Arvidsson and Heyne 2013). The *original model* contains both the gasification and the methanation stages and is considered an excellent base for further analysis of the methanation process. The *original model* is based on the flow sheet shown in Figure 3, a published example of a Tremp methanation process by Haldor Topsøe and information provided by Göteborg Energi.

The $H_2S$ and $CO_2$ removal system are modelled by Arvidsson and Heyne in a separate file using a radfrac absorber model with a built in electrolyte composition called “emdea” commonly used for MDEA equilibrium calculations. This model is used to find an appropriate split ratio for species other than $H_2S$ that may be removed unintentionally by the $H_2S$ removal. (M. Arvidsson, S. Heyne, et al. 2012).

### 3.1.8 OPTIMIZING REACTOR SYSTEM PERFORMANCE

Three main variables are adjusted in the optimization of the process modelling using four main modelling features. The adjusted variables are pressure and temperature of the reactor, as well as the composition of reactants (concentrations). These three variables are altered using compressors, heaters/coolers, condenser units, as well as recycling of product flows. The condensers are implemented to remove $H_2O$, while the recycling is implemented to dilute the reactant stream.

#### *HEAT RECOVERY OPTIMIZATION THROUGH PINCH ANALYSIS*

Aspen has a Pinch analysis tool that can be used to analyse the possibility to integrate the power-to-gas process. The Pinch analysis is a methodology of how to minimize energy consumption of a process by making maximum use of existing heat generated in a process. In the Pinch analysis the process data is represented as a set of streams, as a function of heat load (kW) against temperature (°C). The result is usually presented in a composite curve. The Pinch analysis is used to determine the amount of heat available for district heating in the case of GoBiGas and to optimize the heat utilization at Gasendal.

The result from the Pinch analysis is presented in a grand composite curve. In a grand composite curve all streams are represented in one line. A negative slope of the line represents an excess of heat at that temperature range, a positive slope means that there is a heat demand at that temperature range. If there is an excess of heat at a higher temperature then it can be heat exchanged to a part of the system with a heat demand at a lower temperature. This is graphically shown in the grand composite curve when a line with a negative slope is above a line with positive slope.

## 3.2 ECONOMY

The economic calculations are carried out in the same way for both Gasendal and GoBiGas.

The type of costs and revenues that are associated with the implementation of a power-to-gas process can be divided into four basic categories; investment costs, fixed costs, variable costs and variable revenue. The nature of these costs will decide whether the implementation of a power-to-gas process is economically viable. Investment costs are considered to be initial costs required to implement the power-to-gas process including engineering, equipment and construction costs. Fixed costs are costs that are independent of the production of e-gas at the plant and are not counted in the initial investment costs. These costs include operation & maintenance and fixed water & electricity grid connection costs. The variable costs depend on the e-gas production and include electricity for auxiliary equipment, cooling and other consumed factors. Finally the variable revenues are the income parameters that depend on the e-gas production including the main product, e-gas, but also by-products such as heat and oxygen.

### 3.2.1 ECONOMIC ASSUMPTIONS

The economic lifetimes of the projects are assumed to be 15 years, which is the standard for economic calculation for these types of projects at Göteborg Energi. The economic lifetime should not be confused with the technical lifetime, as the former is the basis for the pay-back period of the project, while the technical lifetime is the actual lifetime of the equipment, which many times can be longer.

The nominal interest rate for the economic calculations is assumed to be 7%, which is also the standard for economic calculations for projects at Göteborg Energi. The nominal interest rate does not take into consideration the inflation rate, but covers the opportunity cost of the investment. The capital recovery factor is used to calculate the annual income required to cover the initial investment required in a project.

The investments costs of different equipment are estimated from contact with manufacturers. The data of the investment costs are for a specific size of the equipment, in cases where the size needed in the process is different from the size stated by the manufacturer, the cost is estimated exponentially with a factor 0,6. Equation (8) shows how a cost adjustment is done for an electrolyser, other equipment use the same equation but change the $H_2$ flow variables.

$$\left(\frac{H_2flow_{actual\ size}}{H_2flow_{size\ at\ which\ price\ is\ known}}\right)^{0,6} = \frac{Price_{actual\ size}}{Price_{size\ at\ which\ price\ is\ known}} \tag{8}$$

# 4 TECHNOLOGICAL ANALYSIS

## 4.1 GENERAL ANALYSIS OF THE SABATIER REACTORS

The following technical analysis presents the different optimizations that can be used to maximize $H_2$ conversion in the process, and to keep the outlet temperature of the reactors below the material constraints set by the catalyst. Finally, a base model for the Sabatier process is presented which will be implemented and modified for the two cases, Gasendal and GoBiGas.

As was stated earlier, the main challenge of utilizing an adiabatic Sabatier reactor is cooling. The temperature can be reduced by adjusting the inlet composition of the reactants entering the reactors, which can be done by dilution, recycling and controlling the ratio of reactants. The outlet temperature of the reactors may not exceed 450°C, due to catalyst material constraints, and the purity of the final product needs to remain above 95%. The three analyses below are carried out by creating a model in Aspen using a Gibbs equilibrium reactor operating at adiabatic conditions. The formation of carbon was monitored in all cases.

### 4.1.1 DILUTION WITH $CH_4$ AND $H_2O$

Dilution with $CH_4$ and $H_2O$ can be easily implemented since a clean $CH_4$ stream is available at Gasendal and $H_2O$ is readily available. The investigation focuses on the addition of $CH_4$ and $H_2O$ in 3 different combinations, a $CH_4$ to $H_2O$ ratio of 1:0, a ratio of 0:1, and an optimal ratio to maximize conversion but prevent carbon formation. The additional dilution stream is added in steps ranging from 0% molar flow of the initial product stream to 100% molar flow of initial product stream. 0% and 100% corresponds to no increase in molecular flow and a doubling of the molecular flow in the reactor respectively. $H_2O$ is added in the form of steam at the reactor inlet temperature. The results are shown in Figure 6 and Figure 7.

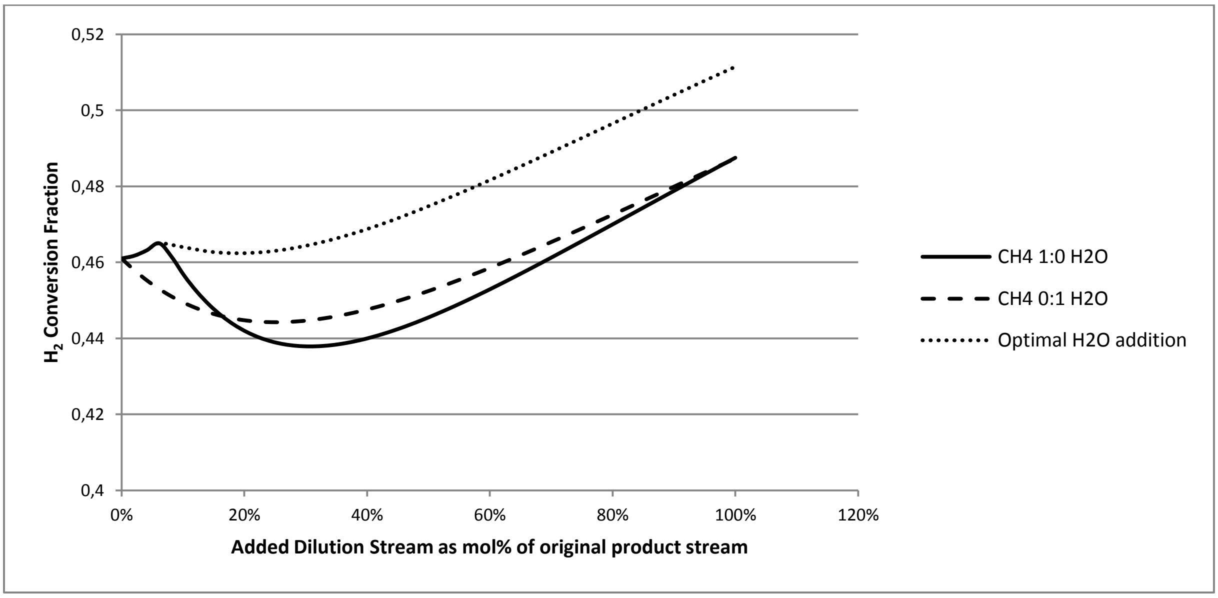


**Figure 6.** $H_2$ conversion fraction as a function of added dilution stream while monitoring carbon formation.

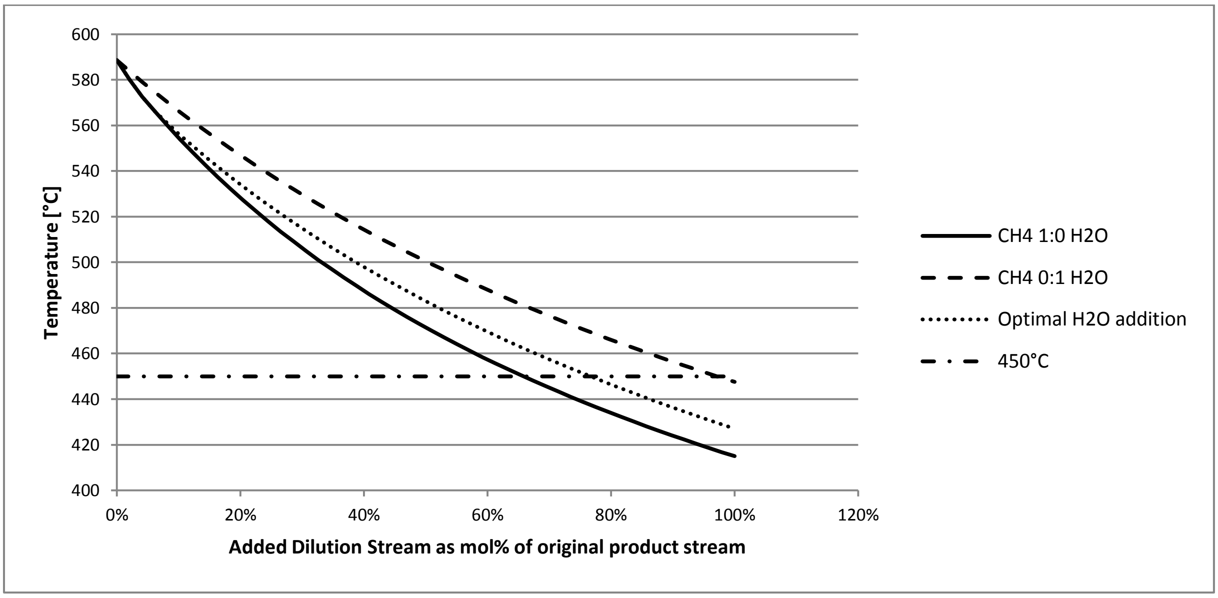


**Figure 7.** Reactor exit temperature as a function of added dilution stream while monitoring carbon formation.

Since the heat capacity of $CH_4$ is approximately 30% higher than the heat capacity of $H_2O$ with respect to volume, the addition of $CH_4$ would provide a better heat sink for the heat of the reaction, which is shown by the steeper gradient of ratio 1:0 than 0:1 in Figure 7. On the other hand, the addition of $CH_4$ aggravates carbon formation, shown by the sudden dip in $H_2$ conversion Figure 6 and therefore a certain amount of $H_2O$ should be added to prevent this. The addition of $H_2O$ has the initial effect of shifting the equilibrium and hence decreasing the $H_2$ conversion, shown by the initial dip of the 0:1 ratio curve in Figure 6. In summary, adding a combination of $CH_4$ and $H_2O$, where the $CH_4$ addition is maximized and the $H_2O$ addition is minimized to prevent carbon formation, optimizes $H_2$ conversion. This situation is shown by the optimal addition curve in Figure 6 and Figure 7.

### 4.1.2 RECYCLING OF PRODUCTS

Another way to dilute the inlet stream is to cool and recycle some part of the product stream. The fraction of recycling was varied between 0 and 0,89, which corresponds to an added dilution stream of 0 and 240 molflow%. The outlet temperature and $H_2$ conversion were monitored, and the result is compared to the optimal $CH_4$/$H_2O$ dilution (investigated in 4.1.1) in Figure 8.

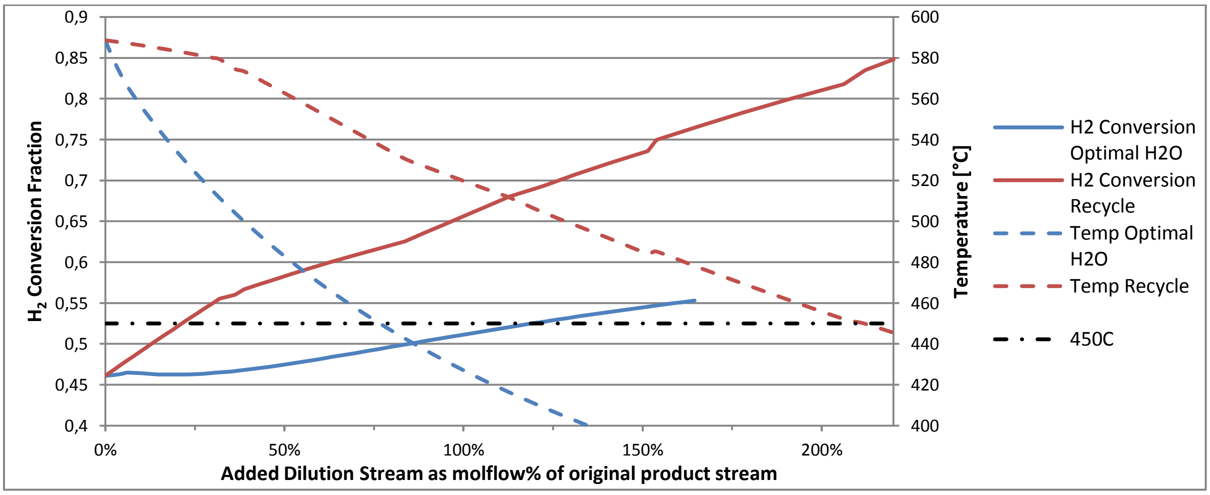


**Figure 8.** $H_2$ conversion fraction and reactor outlet temperature as a function of dilution stream. The figure shows a comparison between dilution with optimal $H_2O$ and recycling of the product stream.

Figure 8 shows that the recycling of product gases is an effective way of increasing the $H_2$ conversion, but is also an inefficient way of cooling the reactor. This stems from the fact that the equilibrium of the reaction is favoured when a mixture of all components is diluted in the reactant stream, as is the case with recycling, as opposed to the dilution with only products. The cooling of the reactor is less efficient due to the relatively high amount of $H_2$ in the recycle stream that lowers the overall heat capacity of the reactant stream. To cool the reactor to 450°C, an addition of 215% of initial product molar flow is required through recycling as opposed to a dilution of 75% with the optimal $CH_4/H_2O$ dilution. On the other hand, at the reactor temperature of 450°C the $H_2$ conversion is 84% for the recycling and only 50% for the optimal $CH_4/H_2O$ dilution.

### 4.1.3 RATIO OF REACTANTS

Another way to cool the reactor is to use under-stoichiometric amount of $CO_2$ or $H_2$ by adjusting the ratio of reactants in the inlet of the reactors. Under-stoichiometric implies that a smaller amount of reactant is provided than the amount that would allow full conversion. The analysis is presented in Figure 9.

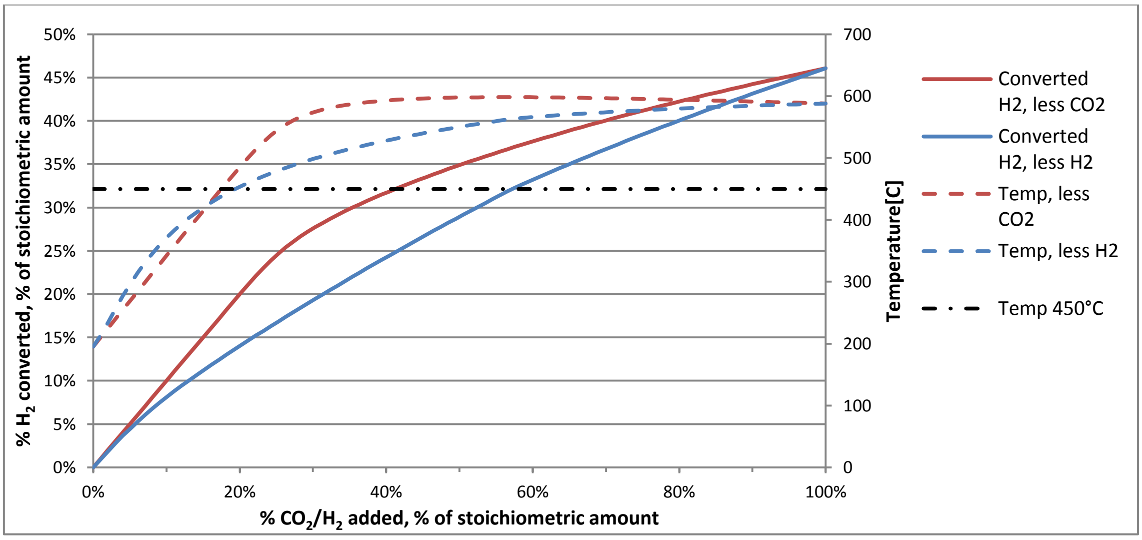


**Figure 9.** The effect of under-stoichiometric addition of reactants on $H_2$ conversion fraction and outlet temperature.

In the case with under stoichiometric $CO_2$ there is no problem with carbon formation but in the case with under stoichiometric $H_2$, carbon formation becomes a problem and $H_2O$ is therefore added to avoid it. It can be seen in Figure 9 that the temperature is increasing rapidly with the amount of $CO_2$ added for the under-stoichiometric $CO_2$ case and reaches the temperature limit of 450°C when 17% of the stoichiometric amount is added. At this temperature the $H_2$ conversion is about 17%. In the case with under-stoichiometric $H_2$, the temperature reaches 450°C when 15% of the stoichiometric amount is added and at this $H_2$ amount the conversion is about 14%. One can therefore conclude that an under-stoichiometric addition of $CO_2$ would be more beneficial with respect to $H_2$ conversion than under-stoichiometric $H_2$. When looking at the heat capacities of $CO_2$ and $H_2$ it seems that it should be the opposite, because of the higher heat capacity of $CO_2$. It must however be noted that because of the 1:4 relation between $CO_2$ and $H_2$ in the Sabatier reaction the molar flow will be higher in the case with under-stoichiometric amount of $CO_2$ and therefore the product stream will be able to absorb more heat. The disadvantage is that it will require a larger reactor.

### 4.1.4 CONCLUSION OF TEMPERATURE CONTROL

From these results the conclusion is that the best option to maximize $H_2$ conversion is to recycle the product gases and only add enough water to avoid carbon formation. This however comes at a cost of a relatively higher volume flow through the reactor than the case of dilution and adjustment of reactant ratio. Using an under-stoichiometric reactant ratio is a way of reducing the reactor size, but also requires more reactors in series to reach full conversion. A base model for the Sabatier reactor system is shown in Figure 10 and the process properties are presented in Table 4.

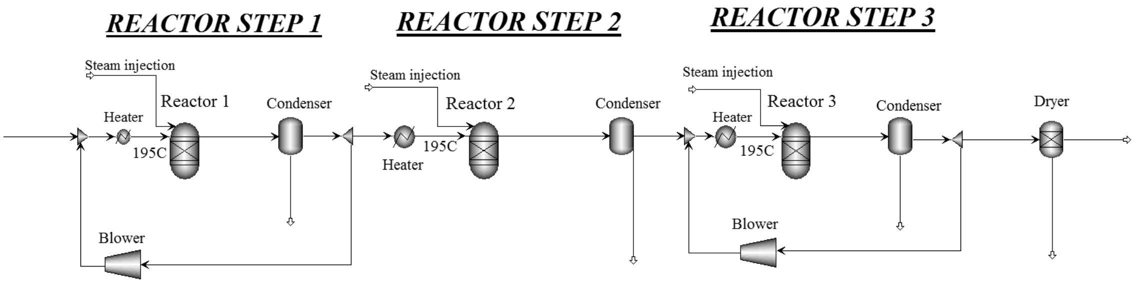


**Figure 10.** Sabatier reactor system design.

**Table 4.** Properties of the reactor system. Composition taken in the inlet and the outlet and out from each reactor step (after the condensers).

| | Inlet | Step 1 | Step 2 | Step 3 | Outlet | Whole system |
|---|---|---|---|---|---|---|
| **% of $H_2$ converted [a]** | - | 88 | 7 | 4 | - | 99,4 |
| **Heat released [b] [kW]** | - | 960 | 70 | 50 | - | 1180 |
| **Recycle fraction** | - | 0,85 | - | 0,75 | - | - |
| **Composition:** | | | | | | |
| **$CH_4$** | 0 | 0,58 | 0,78 | 0,95 | 0,972 | - |
| **$H_2$** | 0,8 | 0,32 | 0,16 | 0,02 | 0,022 | - |
| **$CO_2$** | 0,2 | 0,08 | 0,04 | 0,005 | 0,005 | - |
| **$H_2O$** | 0 | 0,02 | 0,02 | 0,025 | 0 | - |

[a] *% of $H_2$ going into the reactor system,*
[b] *net heat release, including heat for generating the injected steam*

The reactor process utilizes recycling over the first and last reactors to reduce the temperature and maximize $H_2$ conversion. The first reactor is designed to operate below 450°C, while the second and third reactors are designed to maximize $H_2$ conversion and produce a product stream with a $CH_4$ concentration exceeding 95%. The first recycle has a recycle fraction of 0,85 and the last recycle has a fraction of 0,75.

The reactor process utilizes condensation after each reaction step to minimize the gas flow through the recycle and through the reactors themselves. The condensation of $H_2O$ also serves the purpose of minimizing the amount of $H_2O$ entering the reactor and therefore shifts the equilibrium to maximize the $CH_4$ production. To prevent carbon formation, a minimal amount of steam is inserted in to each reactor.

## 4.2 POWER-TO-GAS PROCESS INTEGRATION AT GASENDAL

The $CO_2$ removal induces a heat demand in the upgrading process which is covered by burning natural gas. The natural gas used for steam production in the $CO_2$ removal is around 700kW for the average operation capacity of 1018 $Nm^3$/hr bio-gas (Eklund 2013). Since power-to-gas processes produce excess heat, there is a possibility to utilize this heat to cover the heat demand and thereby save natural gas.

Two alternative designs of the Sabatier reactor system are reached: one design where the Sabatier reactor system is pressurized and one where it operates close to atmospheric pressure. Both recover enough heat to cover the heat demand of the $CO_2$ removal at Gasendal. They also reach a high enough purity of $CH_4$ in the stream out of the system, 96% $CH_4$ or higher. The atmospheric alternative has the benefit of utilizing all available $CO_2$ and the pressurized alternative has the benefit of higher total efficiency but does not use all $CO_2$.

The Aspen model of the Gasendal upgrading process is simplified and only takes into account the heat demand of the $CO_2$ removal. This is because it is the only part of the plant that can be integrated with the power-to-gas process. By using excess heat from the Sabatier reaction to cover the heat demand of the $CO_2$ removal, some natural gas that otherwise would have been used to cover the heat demand can be saved.

The Sabatier reactor is the part of the whole process that is modelled in the greatest detail. The design of the reactor has two counteracting parameters which are optimized; heat recovery and $H_2$ conversion. A reactor system optimized for $H_2$ conversion, Figure 10, is used as the base case. Changes are made on this design to recover enough heat to cover the heat needed in the $CO_2$ removal.

### 4.2.1 SYSTEM OVERVIEW

The parts of the power-to-gas system and the relevant parts of the Gasendal upgrading process are joined together into one model in Aspen. This makes it possible to run a Pinch analysis and to examine the potential of integrating the two processes. Three main parts are included in the model: the $CO_2$ removal at Gasendal, an electrolyser and a Sabatier reactor system.

#### *PLACEMENT*

There are two main alternatives of where to place the Sabatier reactor system in the Gasendal upgrading process. One option is to place the reactor system before the $CO_2$ removal, presented in Figure 11a and the other option is to place it after, Figure 11b.

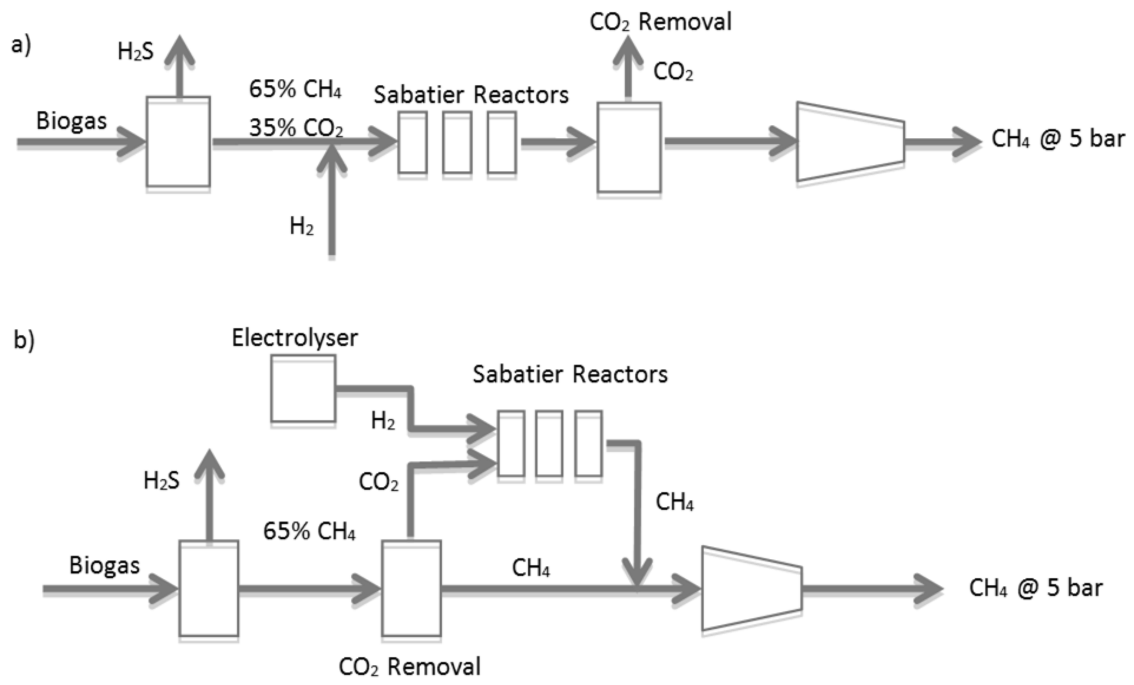


**Figure 11.** a) Reactor placed before $CO_2$ removal. b) Reactor placed after $CO_2$ removal.

In the first option the inlet into the reactor will contain $CO_2$ and $H_2$ as well as $CH_4$. The other option only contains $CO_2$ and $H_2$ into the reactor. The first option has the benefit of making the $CO_2$ removal unnecessary, which saves energy. This option would therefore have more heat available for other uses. The conversion is worse in this alternative due to a higher concentration of products in the reactors, which disfavours the conversion and the volume flows through the reactors are higher. The second option is chosen because of higher $H_2$ conversion and because the $CO_2$ removal unit is already installed. The extra available heat would probably not be possible to use to something useful. There are no other heat demands at Gasendal and the amount of heat is too small to justify a connection to the district heating network. However in a situation where a new upgrading plant is planned this could be worth considering.

### *Atmospheric vs Pressurized Reactor System*

The natural gas distribution network requires a pressure of 6 bar (absolute pressure). To reach this pressure of the e-gas, either the products ($CH_4$) or the reactants ($CO_2$ and $H_2$) have to be compressed. Compressing the reactants will consume more energy and require a more expensive compressor than compressing the e-gas. This is because of the higher volume flow and other difficulties involving compression of hydrogen. On the other hand, it is beneficial for the conversion if the reactants are pressurized as well as for the heat recovery.

It would be beneficial if the existing compressor at Gasendal could be used for compressing the e-gas. However this compressor does not have enough extra capacity to handle the volume flow of the e-gas and an additional compressor is required.

Both the atmospheric and the pressurized alternative have different benefits and therefore both cases are investigated and discussed further in the following sections.

### 4.2.2 INPUT

The heat needed for the $CO_2$ removal was determined from operation measurements to be about 4000 kJ/kg, the details of calculation can be seen in Appendix A - Input Data. The temperature of the outlet stream with $CO_2$ is set to 23°C (Eklund 2013). It is assumed that the flow of raw bio-gas is constant, with a value of 1018 $Nm^3/hr$, the average flow during 2012.

### 4.2.3 RESULTS

Below are the results of the analysis presented. Design 1 is a design where the reactor system operates at close to atmospheric pressure. Design 2 is a design where the reactor system operates at a higher pressure.

#### *DESIGN 1-ATMOSPHERIC REACTOR SYSTEM*

Design 1, the design where the compressor is placed after the reactors, is presented in Figure 12 followed by Table 5 containing the performance of the reactor system.

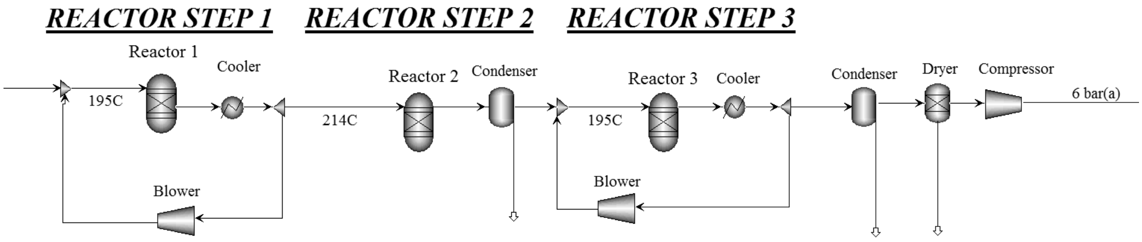


**Figure 12.** The atmospheric reactor system, optimized for heat recovery and $H_2$ conversion.

**Table 5.** The reactor process properties. Composition taken in the inlet and the outlet and out from each reactor step (after the cooler for Step 1 & 3 and after the condenser for Step 2).

| | Inlet | Step 1 | Step 2 | Step 3 | Outlet | Whole system |
|---|---|---|---|---|---|---|
| **% of $H_2$ converted [a]** | - | 80,5 | 10,2 | 8,5 | - | 99,2 |
| **Heat released [b] [kW]** | - | 510 | 450 | 30 | - | 990 |
| **Recycle fraction** | - | 0,76 | - | 0,80 | - | - |
| **Composition:** | | | | | | |
| **$CH_4$** | 0 | 0,23 | 0,60 | 0,75 | 0,962 | - |
| **$H_2$** | 0,8 | 0,23 | 0,25 | 0,02 | 0,030 | - |
| **$CO_2$** | 0,2 | 0,06 | 0,06 | 0,01 | 0,007 | - |
| **$H_2O$** | 0 | 0,48 | 0,09 | 0,22 | 0 | - |

[a] *% of $H_2$ going into the reactor system,*
[b] *note this heat is released in the temperature range of 30-450°C*

The reactor system consists of three reactors, two recycles and two condensers where water is removed. It has been optimized for heat recover as well as $H_2$ conversion.

The $H_2$ conversion is maximized when there is a minimum amount of water in the reaction, just enough water to avoid carbon formation is ideal. However removing water from the stream before a reactor requires heat. Since the stream has to be reheated before entering the reactor to meet the temperature requirement of the catalyst. The recovered heat from cooling the stream, including the condensation of steam, is not released at a high enough temperature to be used to reheat the stream. Therefore there is a net heat requirement when removing water. This is the reason why the water content is only optimized in the last reactor.

The recycling over the first reactor step is implemented to avoid reaching a temperature that would damage the catalyst. The recycle over the last reactor step is used to reach a high enough purity of $CH_4$ in the outlet stream.

The power-to-gas process at Gasendal has been optimized to cover the heat requirement in the $CO_2$ removal unit at Gasendal, because there are no other heat requirements at the plant.

For the heat recovery optimization a Pinch analysis is done. The result from the Pinch analysis of integrating a power-to-gas process in Gasendal is presented in a grand composite curve, Figure 13.

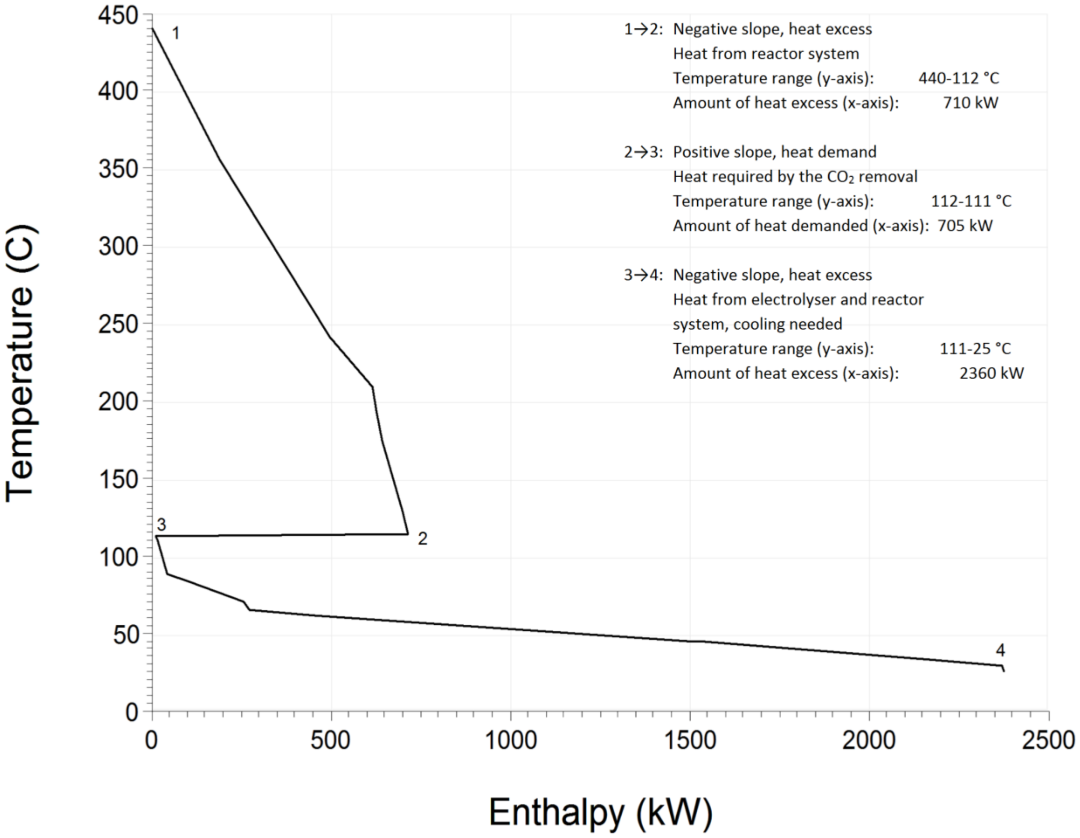


**Figure 13.** The grand composite curve of design 1, showing the optimal process to cover the heat demand of the $CO_2$ removal.

As can be seen in Figure 13 the process generated heat (1→2) is enough to provide for the heat needed in the $CO_2$ removal (2→3).

### *DESIGN 2-PRESSURIZED AND DOWNSIZED REACTOR SYSTEM*

Design 2 (the pressurized alternative) is based on design 1 (the atmospheric alternative) but with some adjustments, an overview can be seen in Figure 14 and the performance of the reactor system in Table 6.

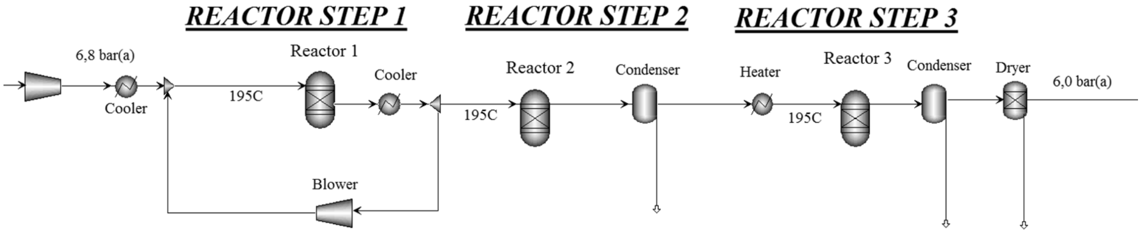


**Figure 14.** The pressurized reactor system, optimized for heat recovery and $H_2$ conversion.

**Table 6.** The reactor Process Properties. Composition taken in the inlet and the outlet and out from each reactor step (after the cooler for Step 1, after the condenser for Step 2 and after reactor for Step 3).

| | Inlet | Step 1 | Step 2 | Step 3 | Outlet | Whole system |
|---|---|---|---|---|---|---|
| **% of $H_2$ converted [a]** | - | 88,4 | 8,5 | 2,5 | - | 99,4 |
| **Heat released [b] [kW]** | - | 610 | 380 | 80 | - | 1070 |
| **Recycle fraction** | - | 0,78 | - | - | - | - |
| **Composition:** | | | | | | |
| **$CH_4$** | 0 | 0,27 | 0,70 | 0,75 | 0,972 | - |
| **$H_2$** | 0,8 | 0,14 | 0,09 | 0,02 | 0,022 | - |
| **$CO_2$** | 0,2 | 0,04 | 0,02 | 0 | 0,006 | - |
| **$H_2O$** | 0 | 0,55 | 0,19 | 0,23 | 0 | - |

[a] *% of $H_2$ going into the reactor system,*
[b] *note this heat is released in the temperature range of 30-450°C*

In design 2 the reaction reaches equilibrium faster than in design 1, due to the higher pressure, this makes it possible to remove the second recycling and still achieve a good $H_2$ conversion. Another advantage with higher pressure is that the condensation of steam occurs at higher temperature which makes it possible to utilize the latent heat of the water in the $CO_2$ removal process. A disadvantage is higher electricity consumption of the compressor.

Because the heat is released at a higher temperature, more heat than required by the $CO_2$ removal will be available with this design. The system is therefore downsized to only release as much heat as is required in the $CO_2$ removal. The advantage of this is that the efficiency will increase while not all available $CO_2$ will be used to produce $CH_4$. The downsized design utilizes about 70% of the available $CO_2$.

### *SUMMARY OF THE TWO DESIGNS*

Design 1(atmospheric) and design 2 (downsized and pressurized) both have different advantages. Properties and the performance for both designs are presented in Table 7 and Table 8.

**Table 7.** Summary of the design differences between design 1 and 2.

| | Design 1 (atm.) | Design 2 (pres.) |
|---|---|---|
| **$H_2$ conversion [$kmole_{H2out}/kmole_{H2in}$]** | 99,2% | 99,4% |
| **$CH_4$ mole fraction out** | 96,2% | 97,2% |
| **Number of recycles** | 2 | 1 |
| **Number of heat exchangers** | 5 | 5 |
| **Number of blowers** | 3 | 1 |
| **Number of compressors** | 1 | 1 |
| **Electrolyser size [$Nm^3_{H2}/hr$]** | 1360 | 940 |
| **% of $CO_2$ flow utilized** | 100% | 69% |

As can be seen design 2 reach a slightly higher $H_2$ conversion and it has a simpler reactor design, with only one recycling. On the other hand it requires a larger and more expensive compressor. Design 2 utilizes only 69% of the $CO_2$ flow and therefore produces 31% less $CH_4$ compared to design 1. Table 8 presents performance data.

**Table 8.** Operation parameters for design 1 and 2 when operating at full capacity.

| Performance | Design 1 (atm.) | Design 2 (pres.) |
|---|---|---|
| **Produced $CH_4$ [$Nm^3$/hr]** | 336 | 233 |
| **Saved NG/Produced $CH_4$ [$kW_{NG}/kW_{CH4prod}$]** | 0,22 | 0,32 |
| **Total electricity consumption [$kW_{EL}$]** | 6644 | 4648 |
| **$El_{compression}/El_{tot}$ [%]** | 1,4 | 2,8 |
| **Efficiency [$kW_{CH4}$/ $kW_{EL}$]** | 47,7% | 47,2% |
| **Total efficiency [$kW_{egas+saved\ NG}$ / $kW_{EL}$]** | 58,3% | 62,3% |
| **Oxygen production [$kg_{O2}$/hr]** | 920 | 630 |
| **Water consumption [$m^3_{H2O}$/hr]** | 1,2 | 0,8 |
| [a] *$kW_{CH4}$ based on LHV of $CH_4$ 802,3 kJ/$mol_{CH4}$* | | |

Both designs are able to cover the heat demand of the $CO_2$ removal, 705 kW. The electricty consumption related to compression is higher for design 2, indicating the high amount of energy required to compress hydrogen. The e-gas ($CH_4$) production is higher for design 1 becasue it utilizes all available $CO_2$. On the other hand, the total efficiency is about 4 percent points higher for design 2. This is becasue the saved natural gas contributes with a larger share to the total efficiency for design 2 compared to design 1.

### 4.2.4 OPERATION

The idea is to run the process when the electricity price is low and turn it off when it is high. Because of fluctuations in the electricity price the process will have to be able to be turned off and on. However, completely shutting down the Sabatier reactor is rather complicated. It would require the reactor to be cooled down with nitrogen and then be sealed in a slightly pressurized nitrogen atmosphere for as long as the reactor is not running. Before operation the reactor would also have to be preheated with nitrogen. This is all done to protect the catalyst from coming into contact with oxygen (Culmsee 2013). An alternative way of avoiding catalyst contact with oxygen would be to lower the flow through the reactor but never shutting it down completely.

During operation the electrolyser can easily vary its $H_2$ production between 20 and 100% of its capacity. A change from production at 100% capacity to 20% takes about 10 minutes (Taalesen 2013). However shutting down the electrolyser completely and then start it up to full capacity takes several hours (Taalesen 2013).

There are no known problems with reducing flow through the reactor (Seemann 2013), however it must be investigated further to confirm that that is the case. In this project it is assumed that there are no problems associated with going down in capacity to 20% of the designed flow through the reactor. This would allow for a simple way to deal with periods when the electricity price is high and it is not economical to run the process. This would also reduce the need for $N_2$ addition during production stoppage periods.

There will be a cost associated with adjusting the process after the electricity price, no matter how it is done. It is therefore of interest to take this into account in the economical calculations and to determine how large it is. In a real case there might be periods when it is a good idea to shut down the plant completely and other periods when it is better to run at 20%. However to simplify the calculations it is assumed that the process is never completely shut down and that it is running at 20% during the periods when it is not profitable to run the process. The performance of the process is shown in Table 9.

**Table 9.** Operation parameters for design 1 and 2 when operating at 20% of capacity operation.

| Performance at 20% operation | Design 1 (atm.) | Design 2 (pres.) |
|---|---|---|
| **Produced $CH_4$ [$Nm^3$/hr]** | 67 | 47 |
| **Saved NG/Produced $CH_4$ [$kW_{NG}/kW_{CH4prod}$]** | 0,22 | 0,32 |
| **Total electricity consumption [$kW_{EL}$]** | 1193 | 836 |
| **$El_{compression}/El_{tot}$ [%]** | 1,6 | 3,1 |
| **Efficiency [$kW_{CH4}$/ $kW_{EL}$]** | 53,1% | 52,5% |
| **Total efficiency [$kW_{egas+saved\ NG}$ / $kW_{EL}$]** | 65,0% | 69,4% |
| **Oxygen production [$kg_{O2}$/hr]** | 180 | 130 |
| **Water consumption [$m^3_{H2O}$/hr]** | 0,24 | 0,17 |
| [a] *$kW_{CH4}$ based on LHV of $CH_4$ 802,3 kJ/$mol_{CH4}$* | | |

The efficiency increase of the process when it is operated at 20% is due to an increase in electrolyser efficiency as a result of lower current densities.

## 4.3 POWER-TO-GAS PROCESS INTEGRATION AT GOBIGAS

The analysis will focus on the methanation stage of the GoBiGas process, as this is where $H_2$ and CO is reacted to form $CH_4$ and where the separation of excess $CO_2$ takes place. The methanation reaction releases heat, but since the GoBiGas plant in its initial design already produces an excess amount of heat during the gasification and methanation stages, there is no incentive to investigate the potential of utilizing the heat produced from an increased methanation in the gasification process. Therefore, only the methanation stage will be treated and no evaluation will be made on the gasification stage. The analysis in this section will focus on evaluating the effect of adding $H_2$ to the system, and will determine which limiting factors exist for the system.

### 4.3.1 SYSTEM OVERVIEW

The parts of the power-to-gas system and the relevant parts of the GoBiGas bio-gas production plant are joined together into one model in Aspen. This allows the integration of the Sabatier reactor system in the existing GoBiGas process to be examined.

The two options considered in this analysis are either the addition of $H_2$ in the main syngas stream prior to methanation, shown in Figure 15a below, or the separate implementation of the Sabatier reactor system parallel to the existing methanation shown in Figure 15b.

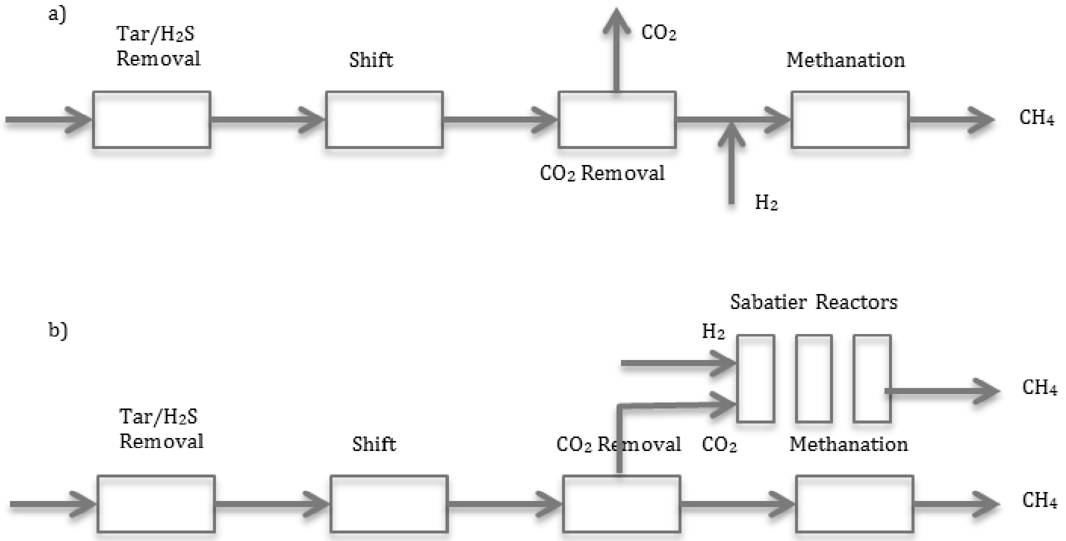


**Figure 15.** Options for $H_2$ addition in GoBiGas process integration.

### 4.3.2 INPUT

The composition of the syngas originating from the gasification and entering the compressor prior to the methanation stage was confirmed by Göteborg Energi and is shown in Table 10. The total syngas flow is assumed to be 6266 $Nm^3$/hr.

**Table 10.** Composition of input syngas.

| Compound | Molar fraction |
|---|---|
| **$H_2$** | 0,38 |
| **CO** | 0,21 |
| **$CO_2$** | 0,21 |
| **$CH_4$** | 0,10 |
| **Various alkenes** | 0,045 |
| **$H_2O$** | 0,03 |
| **$H_2S$** | 0,02 |
| **$N_2$** | 0,01 |

Out of the approximately 45 kmol/hr (1115 $Nm^3$/hr) of $CO_2$ separated from the process in the 20$MW_{SNG}$ plant, only 9 kmol/hr (225 $Nm^3$/hr) is considered to be available for methanation due to the use of the $CO_2$ in other parts of the process (Gunnarsson 2013).

The initial pressure of the process is set to 16bar in the model used for this analysis. This change in pressure from 30bar in the *original model* decreases the CO conversion in the methanation reactors and requires the addition of a 4th methanation reactor in the Tremp methanation process to achieve the final $CH_4$ purity of 95%. The use of four adiabatic methanation reactors in series has been documented in tests done by DNV KEMA and Haldor Topsøe and the modification is there for accepted as an appropriate measure to increase final $CH_4$ purity. The model used in this analysis will from now on be called the *modified model* and is shown in Figure 16.

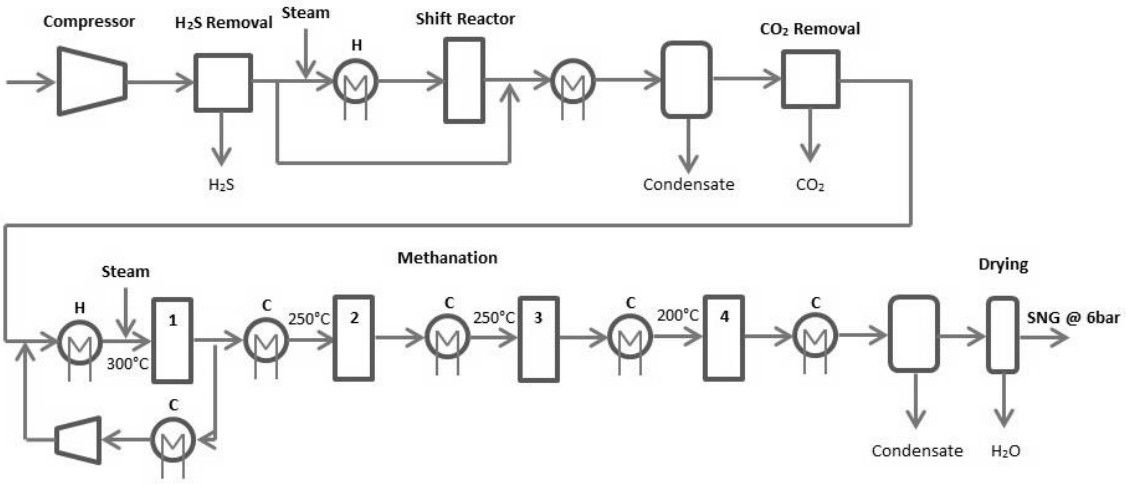


**Figure 16.** Flow sheet of modified GoBiGas methanation process.

### *OPTION 1 – ADDITION IN SYNGAS STREAM*

The analysis is carried out in Aspen where the $H_2$ is added prior to the methanation reactor system. This $H_2$ addition requires an adjustment of the shift reactor to achieve a CO:$H_2$ ratio of 1:3. The added $H_2$ is varied between 0 and 120 kmol/hr corresponding to an electrolyser of rating 0 – 13,7 MW.

### $CO_2$ REMOVAL

The results show that as the amount of added $H_2$ increases, the amount of $CO_2$ that is removed after the shift reactor decreases. This is due to the fact that as more $H_2$ is added, the ratio of $H_2$ to CO increases and therefore a smaller fraction of the stream needs to be reacted in the shift reaction hence decreasing the amount of produced $CO_2$. In this application, only 20% of the $CO_2$ is considered available since 80% of the stream is dedicated to various processes on site. The dotted black line in Figure 17 illustrates this constraint. This corresponds to a $H_2$ addition of 37 kmol/hr to the process stream. The units are expressed as the ratio between added $H_2$ and required $H_2$, where the required $H_2$ represents the amount of $H_2$ added to completely bypass the shift reactor, which is 95 kmol/hr. In this situation, only the original $CO_2$ from the syngas stream is removed.

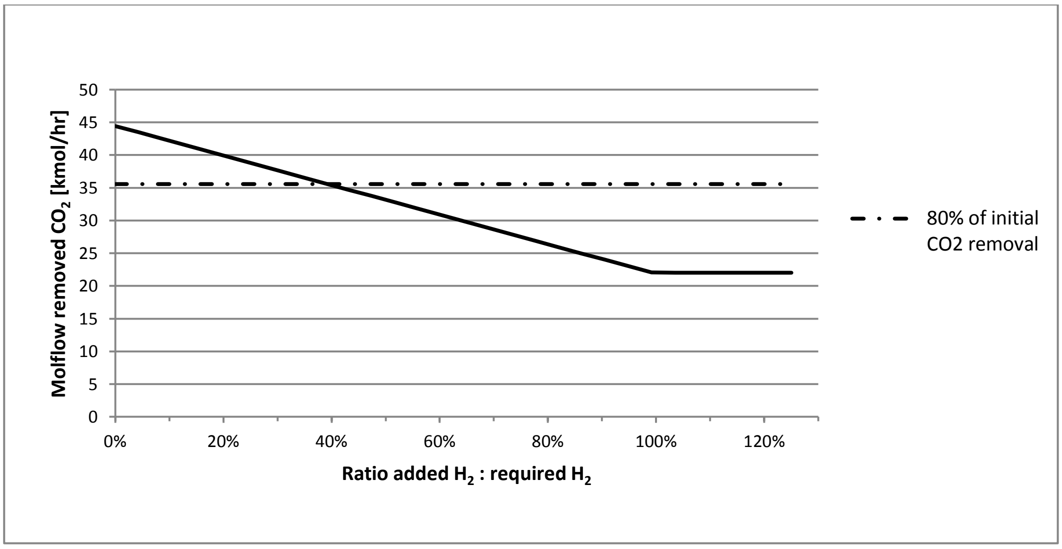


**Figure 17.** Removed $CO_2$ as a function of added $H_2$.

### REACTOR OUTLET TEMPERATURE

It is important to investigate the scenario where there is no constraint on the amount of available $CO_2$, as the fraction of available $CO_2$ is likely to increase with an increase in plant size. Therefore, the exit temperatures of the first and second reactor are monitored with the addition of $H_2$. The reactor outlet temperature of the first and second methanation reactors are constrained to 650°C and 450°C due to the temperature limit of the Haldor Topsøe MCR and PK-7R catalyst respectively. The effect of added hydrogen can be seen in Figure 18.

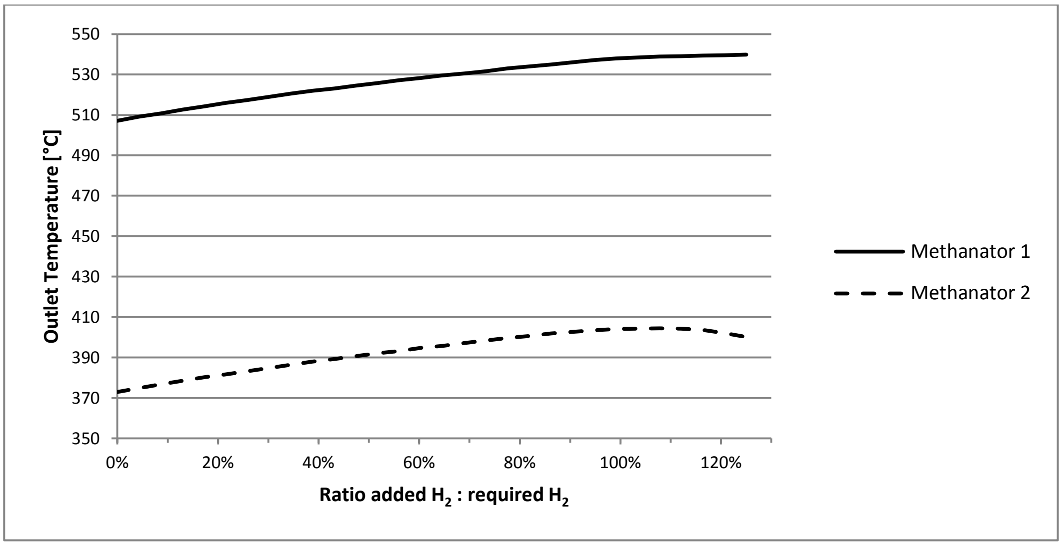


**Figure 18.** Outlet temperature of methanator 1 & 2 as a function of added $H_2$.

The results show that the exit temperatures of the two reactors increase with added $H_2$, but never exceed the limiting temperatures of the two catalysts. The increase in temperature is due to a change in the composition of the reactant gas at the inlet of the reactor. In the inlet streams of both reactors, the $H_2$ and CO concentrations increase even though the ratio of $H_2$ to CO remains at approximately 3:1 at the inlet of the first methanation reactor. The rather low temperature increase is the result of the cooling effect of the recycle stream. The omission of the recycle stream would result in a far larger outlet temperature of the methanation reactors.

### OTHER ASPECTS AFFECTED BY $H_2$ ADDITION

There are numerous other factors that are affected by the addition of $H_2$. The first is the fraction of the process stream that is bypassed around the shift reactor. Recall that the reason for the shift reactor in the first place is to boost the $H_2$ concentration to a $H_2$ to CO ratio of 3:1 prior to the methanation. An increase in initial $H_2$ concentration will decrease the need for shifting and consequently increase the fraction of the stream that bypasses the shift reactor. The analysis shows that the stream entering the shift reactor diminishes until the reactor is completely bypassed at a $H_2$ addition of 95 kmol/hr. The increasing fraction of bypass may require a redesign in the pipes and valves transporting the bypassed stream.

The second factor is the $CH_4$ concentration of the product stream prior to connection to the natural gas grid. As a result of the increased concentration of $H_2$ in the reactant stream, the outlet temperatures of all methanation reactors increased slightly. Despite this increase in exit temperature of the reactors, the $CH_4$ concentration increases slightly as well, which can be seen in Figure 19. The sharp dip after approximately 95 kmol/hr is the result of the completely bypassed shift and the consequent excess $H_2$ in the reactant stream.

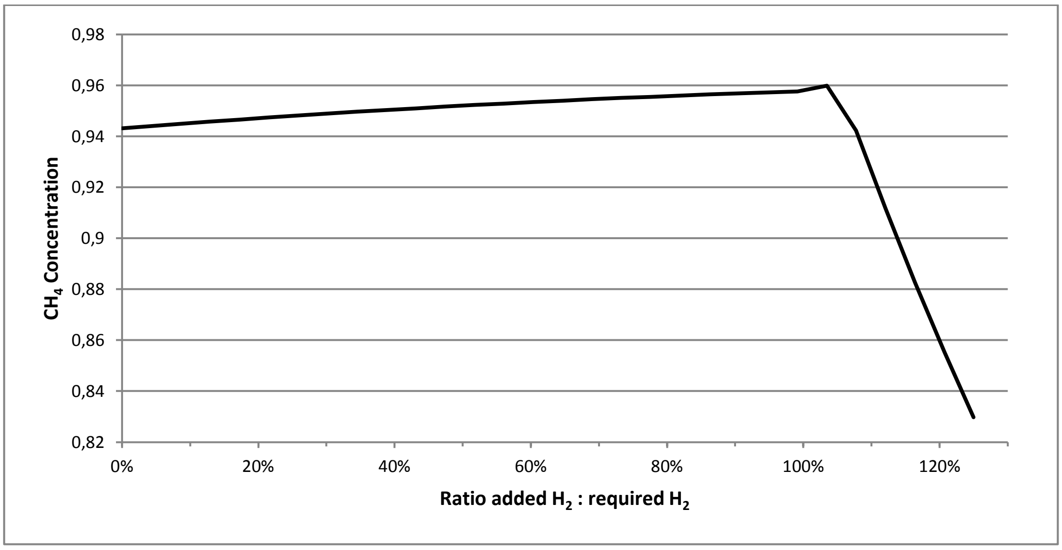


**Figure 19.** $CH_4$ outlet concentration as a function of added $H_2$.

This stems in the fact that as the $H_2$ is added, the composition of the reactant gases entering the methanation reactors changes, resulting in a slightly different equilibrium. The overall effect of the addition of hydrogen is that the equilibrium is shifted towards products, thus counteracting the slight increase in outlet temperature.

Another aspect that has been monitored while increasing the added amount of $H_2$ is the increased compressor work. This aspect is important both with regards to increased energy consumption of the process but also with respect to the physical limits of the existing compressor. The increase in compressor work with an increase in added $H_2$ is linear and the physical limit in increased volume flow in the compressor is assumed to be 20% higher than the original flow. (Anderson 2013) This limit is reached after a $H_2$ addition of 50 kmol/hr.

The addition of $H_2$ in the existing methanation reactor system reduces the amount of $CO_2$ removed in the $CO_2$ removal system by 20%. This $CO_2$ that no longer is removed results in an energy saving in the $CO_2$ removal system that amounts to 25% of the energy required for $CO_2$ removal, which is considered to be 4022 kJ/kg$CO_2$. Only 25% can be accounted as energy savings since the $CO_2$ removal system currently recovers 75% of the heat used for $CO_2$ removal (Arvidsson and Heyne 2013). The heat savings amount to 114kW.

### OVERVIEW OF OPTION 1

The change in shift and methanation reactor heat duties with the addition of $H_2$ can be seen in Table 11. The steam injection into the shift reactor decreased by 60% from 87,8 kmol/hr with no $H_2$ addition to 55 kmol/hr with $H_2$ addition.

**Table 11.** Change in heat release with $H_2$ addition.

| *Heat Duty* | *Change in heat release [kW]* |
|---|---|
| **Shift Reactor** | -590 |
| **1st Methanation Reactor** | 1261 |
| **2nd Methanation Reactor** | 73 |
| **3rd Methanation Reactor** | 42 |
| **4th Methanation Reactor** | 235 |

The decrease in both steam injection and heat release in the shift reactor as a result of $H_2$ addition is due to the increase in bypass fraction around the shift reactor.

The specifications of the process design are shown in Table 12 for the process comparing the scenario where no $H_2$ is added and the scenario when 37 kmol/hr $H_2$ is added to get an overview of the effect of adding $H_2$.

**Table 12.** Specification summary of option 1.

| *Parameter* | *No $H_2$ addition* | *37 kmol/hr addition* |
|---|---|---|
| **$CO_2$ removed [kmol/hr]** | 45,692 | 36,41 |
| **Total compressor work [kW]** | 885,75 | 1013,45 |
| **Recycle fraction [ ]** | 0,49 | 0,68 |
| **1st methanator outlet temp [°C]** | 507 | 522 |
| **Final outlet $CH_4$ purity [ ]** | 0,954 | 0,955 |
| **Total available heat duty [kW]** | 2550 | 3738 |
| **$H_2$ Conversion Efficiency** | - | 0,992 |
| **Electricity-e-gas Efficiency** | - | 0,470 |

### *OPTION 2 – SEPARATE IMPLEMENTATION OF SABATIER REACTOR SYSTEM*

In the second option for $H_2$ addition, the $CO_2$ is separated from the process stream and combined with $H_2$ prior to methanation reactors. The main difference between the option where $H_2$ is injected in the process stream prior to the shift reactor and the second option being described is the installation of the separate methanation process. The same limitation applies for this option as the first option, where the amount of available $CO_2$ restricts the maximum $H_2$ addition to 37 kmol/hr. This option of using a separate methanation process is identical to the situation described for the Gasendal plant earlier and a short recap of the most important results will be described below. The decisive aspect when analysing option 1 and 2 is which option is the most economical. The proposed design of the process design is shown in Figure 14.

The calculated $CH_4$ purity of the Gasendal reactor design is 96,2%, which is above the requirement for the GoBiGas plant. The amount of recycle around the 3rd reactor can be minimized to obtain a $CH_4$ purity of 95% and hence minimize the investment cost of the 3rd reactor, the blower and also the power consumption of the blower. The process produces an excess of 835 kW of heat. The specifications of the process are summarized in Table 13.

**Table 13.** Specification summary of option 2.

| *Parameter* | *No $H_2$ addition* | *37 kmol/hr addition* |
|---|---|---|
| **$CO_2$ removed [kmol/hr]** | 45,692 | 36,412 |
| **Total compressor work [kW]** | 885,75 | 946,22 |
| **Recycle fraction [ ]** | 0,49 | 0,49 |
| **1st methanator outlet temp [°C]** | 507 | 507 |
| **Final outlet $CH_4$ purity [ ]** | 0,954 | 0,953 |
| **Total available heat duty [kW]** | 2550 | 3385 |
| **$H_2$ Conversion Efficiency** | - | 0,994 |
| **Electricity-e-gas Efficiency** | - | 0,472 |

### *CONCLUSION AND DISCUSSION*

The results of the technical sensitivity analysis show that the limiting factors of the hydrogen addition are the $CO_2$ removal and the compressor volume flow capacity. The $CO_2$ removal allows a reduction in $CO_2$ of 20%, which allows a $H_2$ addition of 37 kmol/hr, while the compressor allows a volume flow increase of 20% and therefore allows a $H_2$ addition of 50 kmol/hr. Hence, without any modifications to the GoBiGas process design, an $H_2$ addition of 37 mol/hr can be made which corresponds to a boost in $CH_4$ production of 11% (2,2MW).

The first option does not require the investment of new Sabatier reactors since it utilizes the existing methanation reactors. The introduction of $H_2$ in the process stream will alter the operation of the shift and methanation reactors. A larger fraction of the main stream will bypass the shift reactor thus decreasing the heat release in this unit. The overall heat release will still increase due to a larger heat release in the methanation reactors. Another intrinsic effect of introducing $H_2$ in the process stream is the decrease in removed $CO_2$ and a resulting decrease in heat demand in the $CO_2$ removal unit. The second option requires the investment of new Sabatier reactors but allows the Sabatier reactors to be designed in an optimal way and has no effect on the existing methanation process.

As has been explained earlier, the model used is based on an existing model created by Maria Arvidsson and Stefan Heyne. The validity of this model can be questioned, especially considering the confidentiality of the methanation process design of GoBiGas. The model used is the result of discussions with people working in close collaboration with GoBiGas and the use of published solutions from Haldor Topsøe, for example the Tremp solution. On the other hand, Haldor Topsøe clearly state that they customize solutions for their customers and we have been informed that the exact specifications of the GoBiGas plant have not been published and are therefore not used in this model. The accuracy of the model is however considered to be sufficient to analyze the effect of adding H2 and the conclusions are considered valid.

# 5 Economic Analysis

## 5.1 Economy of Integrating a Power-to-Gas Process at Gasendal

### 5.1.1 Input

The different costs and prices are presented in the economic background, Appendix A - Input Data. The investment costs for the equipment presented in the economic background are not for the size corresponding to design 1 and 2, and are therefore adjusted to the right size using equation (8). In Table 14 are these adjusted investment costs presented for design 1 and 2.

**Table 14.** Investment costs for the scale that is used at Gasendal. The size of each equipment is also shown.

| Equipment | Design 1 (atm.) Cost [MSEK] | Design 2 (pres.) Cost [MSEK] | Design 1 (atm.) size | Design 2 (pres.) size | |
|---|---|---|---|---|---|
| **Electrolyser** | 35,8 | 24,5 | 1400 | 940 | [Nm$^3_{H2}$/hr] |
| **Sabatier reactor** | 13,2 | 10,6 | 350 | 235 | [Nm$^3_{CH4}$/hr] |
| **Dryer** | 0,2 | 0,2 | 350 | 235 | [Nm$^3$/hr] |
| **Compressor** | 0,8 | 2,9 | 350 | 1175 | [Nm$^3$/hr] |
| **Total cost** | 50 | 38,2 | - | - | |

The total investment cost is estimated to 50 MSEK for design 1 and 38 MSEK for design 2, where the difference is mainly due to the investment cost of the electrolyser. The input data for the economic analysis of integrating a power-to-gas process at Gasendal is presented in Table 15.

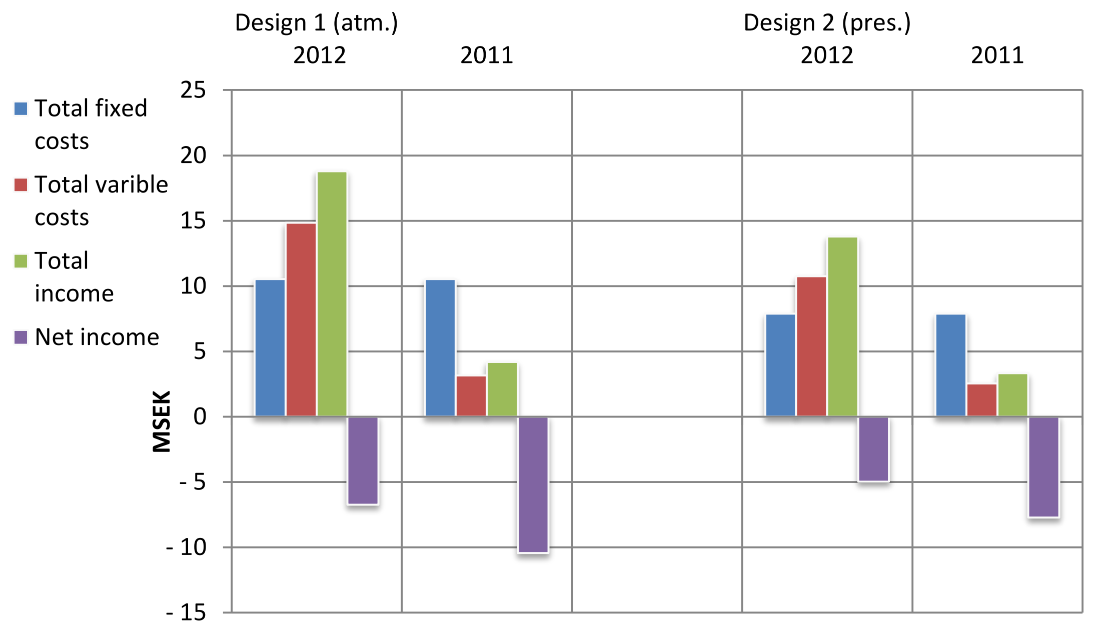
**Table 15.** Input data to the economical calculations. Amore detailed description of how the different inputs were obtained can be found in Appendix A - Input Data.

| Input, costs and prices | | Design 1 | Design 2 |
|---|---|---|---|
| **Investment cost** | SEK | 50 000 000 | 38 200 000 |
| **Interest** | % | 7 | 7 |
| **Economical life time** | Years | 15 | 15 |
| **Electricity price** | SEK/kWh$_{el}$ | Hourly prices 2011 & 2012 | |
| **Electricity distribution, fixed** | SEK/year | 2 528 205 | 1 771 322 |
| **Electricity distribution, variable** | SEK/kWh$_{el}$ | 0,031 | 0,031 |
| **Gas distribution** | SEK/kWh$_{SNG}$ | 0,1 | 0,1 |
| **Operation & Maintenance** | SEK/year | 2 500 000 | 1 910 000 |
| **Water fixed** | SEK/year | 4600 | 4600 |
| **Water variable** | SEK/m$^3_{H2O}$ | 4,8 | 4,8 |
| **Price SNG** | SEK/kWh$_{SNG}$ | 0,7 | 0,7 |
| **Price $O_2$** | SEK/kg$_{O2}$ | 0,325 | 0,325 |
| **Price NG** | SEK/kWh$_{NG}$ | 0,25 | 0,25 |

The process is run at full capacity when the running costs are lower than the income from selling the methane and the oxygen. When the running costs are higher than the income the process is run at 20%. The electricity price used changes each hour and result in a different net income each hour of production. For the power-to-gas process at the Gasendal plant the income comes from selling the produced e-gas, saving natural gas by using heat from the process in the $CO_2$ removal and by selling oxygen that is being produced in the electrolyser. The costs consist of fixed and variable costs. The fixed costs are: investment cost, operation and maintenance cost, fixed electricity distribution cost and the water fixed cost. The variable running costs consist of: electricity, electricity distribution, natural gas distribution and water. Cost of operation at 20% during periods with high electricity prices are also taken into account. Note that the electricity price used is historical data coming from Nordpool spotmarket and can be seen in Figure A1 in Appendix A. One can see that the electricity price is generally lower 2012 than 2011, the average price was 0,28 SEK/kWh for 2012 compared to 0,43 SEK/kWh for 2011.

### 5.1.2 RESULTS

The economic analysis of integrating a power-to-gas process at Gasendal shows clearly that it is not economical to invest in this kind of a plant today. A summary of the results is presented in Figure 20.



| | | | | | |
|---|---|---|---|---|---|
| Operation[hr/yr]: | 6 977 | 1 557 | | 7 190 | 1 738 |
| Annual income[MSEK]: | -6,6 | -9,5 | | -4,9 | -7,1 |
| e-gas prod [GWh/yr]: | 23 | 9 | | 16 | 7 |

**Figure 20.** Overview of the economics for design 1 and 2 when running process with electricity prices from 2011 and 2012. The number of operating hours, the annual income and yearly e-gas production for 2011 and 2012 is shown below the chart.

As can be seen the net income is negative for both designs and for electricity prices from 2011 and from 2012. Design 2 has under these circumstances better net income than design 2, even if design 1 has a larger total income, this is because of the lower fixed costs for design 2.

The electricity price has an impact on the number of hours when it is profitable to run the process. With the electricity prices of 2011 it is only profitable to run the process about 1500 hours compared to prices of 2012, for which it is profitable to run the process for about 7000 hours. This also has a large impact on the total income, which is much lower for 2011 compared to 2012.

The total e-gas production is also affected by the electricity price; design 1 with electricity prices for 2012 has the highest production, about 23 GWh of e-gas is produced, that is about one third of the current production at Gasendal.

Figure 21 shows how large share of the income that is coming from each income source. For the power-to-gas process at Gasendal there are three income sources: e-gas produced, saved NG and oxygen produced.

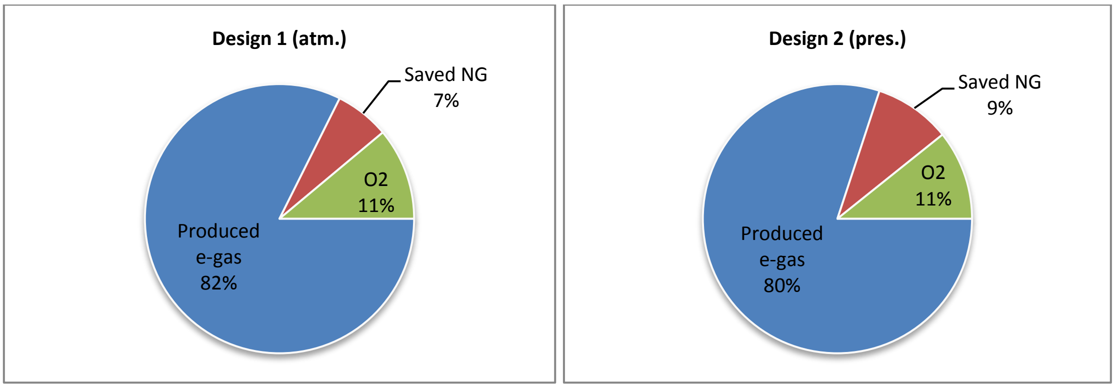


**Figure 21.** Shows how the income is distributed between the produced e-gas, the saved NG and the oxygen production for design 1 and 2.

In both designs the produced e-gas is the main income source, however the other two incomes, saved natural gas and produced $O_2$, contributes with a considerable amount and have a positive effect on the overall economy of the process. Design 1 get a larger portion of its income from the produces e-gas compared to design 2, where the saved NG stands for a 2 % larger share of the income. It is beneficial for the overall economy of the process if the share of the produced e-gas is small, because that means the secondary flows are better utilized. It should be noted that if the NG would have the same value as the produced e-gas then the share of the saved NG would be larger and design 2 would benefit more. This could be the case in a future where it is very expensive to emit $CO_2$.

Figure 22 shows the cost distribution with electricity prices from 2012 and 2011. The cost distribution is almost looking the same for both designs there are more differences between 2011 and 2012, therefore is Figure 22 only showing the cost distribution of design 1.

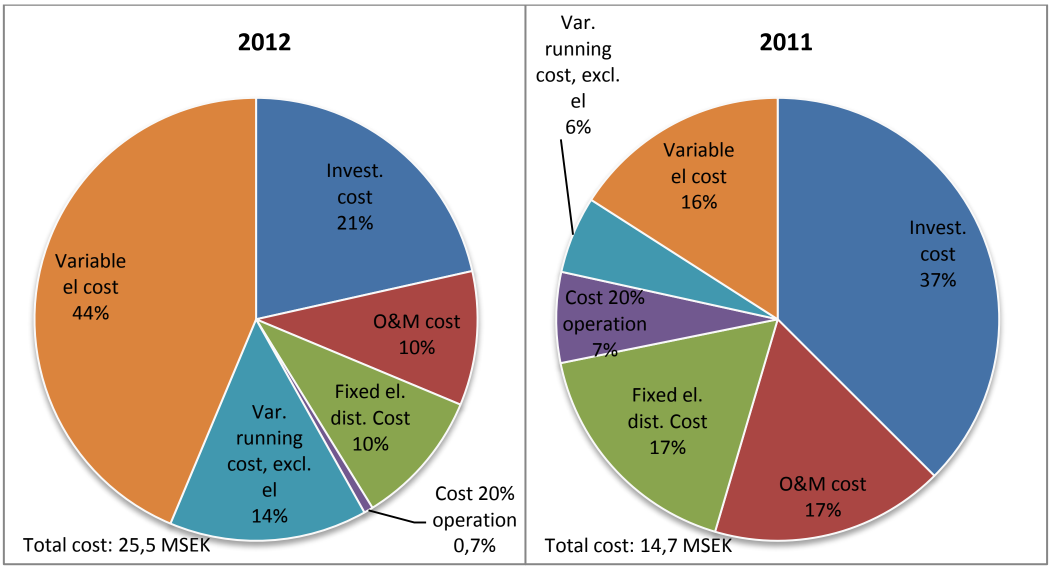


**Figure 22.** The distribution of the costs for the electricity prices during 2012 and 2011 for design 1.

The cost distribution is highly effected by the electricity price and how many hours it is running per year. During 2012 when the process is running almost continuously the variable costs stand for a large share of the total cost and especially the electricity cost which stand for 44% of the total cost. During 2011 the electricity cost stands for only 16%. Of the fixed costs the investment cost stands for about half of the total fixed costs, and the cost for O&M and electricity distribution fixed cost stand together for the other half.

The cost of running the process at 20% of its capacity is here shown to be of significance during a year with high electricity prices. For 2011 about 7% of the total costs but only 1% for 2012, this is explained by the few hours the process is running at 20% for 2012. During 2011 it account for about 1 MSEK and it might be worth considering shutting down the plant completely during periods when high prices are expected, for example during cold winter days.

### 5.1.3 SENSITIVITY ANALYSIS

A sensitivity analysis is made to show how changes in different parameters affect the net income of the process. This is done for electricity prices for 2012 and 2011 and for the parameters: SNG price, investment cost, electricity price and efficiency. The sensitivity is done on the electricity prices for 2011 and 2012 because they show the price each hour of the year. When a parameter is changed that could affect whether or not to run the plant for a specific hour and therefore have an effect on the result. If just an average electricity price would have been used then this aspect of whether or not the plant is run at full capacity would not have been possible to account for.

#### *SNG PRICE*

How changes in the SNG price affect the profitability of the process is shown in Figure 23. The reference in this analysis is 0,7 SEK/kWh$_{SNG}$, a price believed to be a reasonable assumption within the relevant future (Zinn 2013).

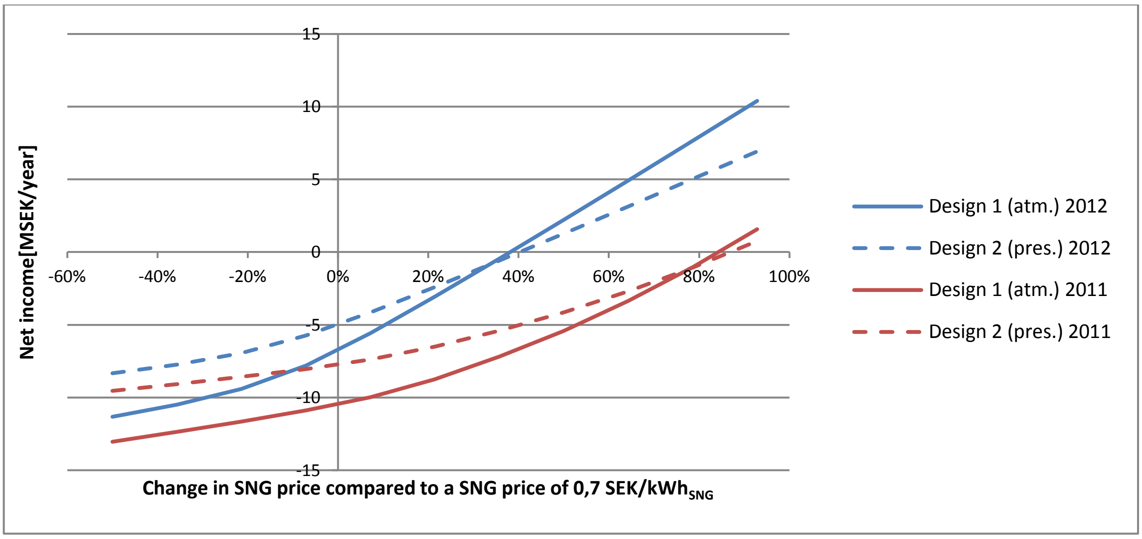


**Figure 23.** Sensitivity analysis of how changes in SNG price affect the net income of the process.

Both designs are compared in Figure 23 to determine which one becomes profitable first. The change in the slope of the curves is due to changes in the number of hours the plant is operated at full capacity. When the SNG price is low then the plant is not operated as often as when the price is higher.

Design 2 is the better choice before the breakeven point but after the breakeven point design 1 will generate more net income. This occurs because of the lower fixed costs, higher total efficiency and lower $CO_2$ utilization of design 2. From this analysis one can conclude that design 1 is the best choice and that the SNG value has to increase 40-80% before the process becomes profitable.

For the following sensitivity analyses only shows the results of design 1 because the same pattern as is shown in Figure 23 holds true for the other sensitivity analyses as well. The pattern that design 1 reaches profitability first and then has higher net income than design 2.

### *Investment Cost*

Figure 24 shows how changes in the investment cost affect the net income of the plant for electricity prices from 2011 and 2012. The reference point is the value estimated for investing in the process today, 50 MSEK. A change in the in the investment cost only changes the annual fixed cost therefore the lines are straight.

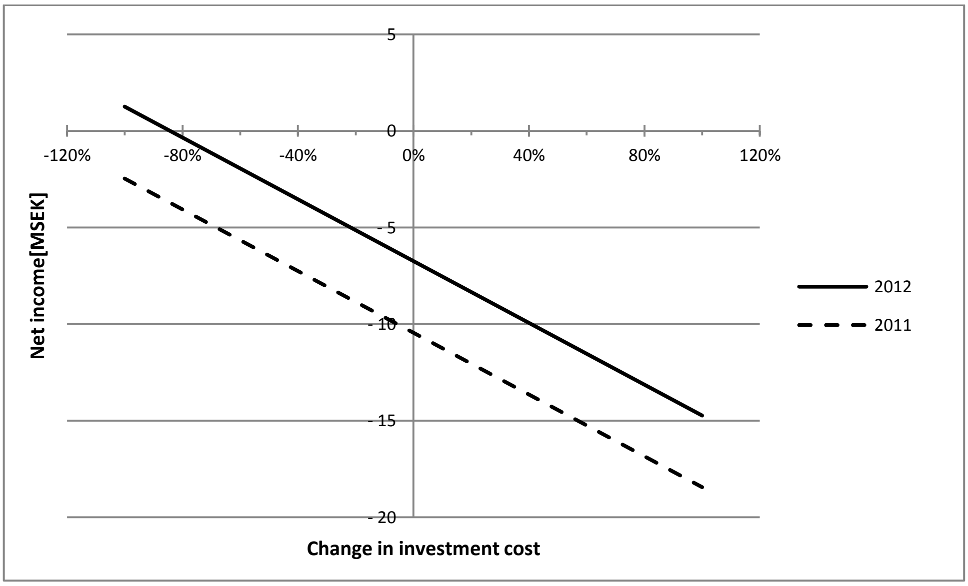


**Figure 24.** Sensitivity analysis of how changes in investment cost affect the net profit of the process.

It shows that lowering the investment cost have a positive effect on the net income, if the investment would be 80% less than today then the plant would be profitable for 2012. During 2011 the plant would not have been profitable even if the investment cost would have been zero. A lower investment cost alone would therefore not be enough to make the process worth investing in.

### *Electricity Price*

Figure 25 shows the effect of changing the electricity price. The electricity price has been changed with X % for each hour of the year, to keep fluctuations in the electricity price.

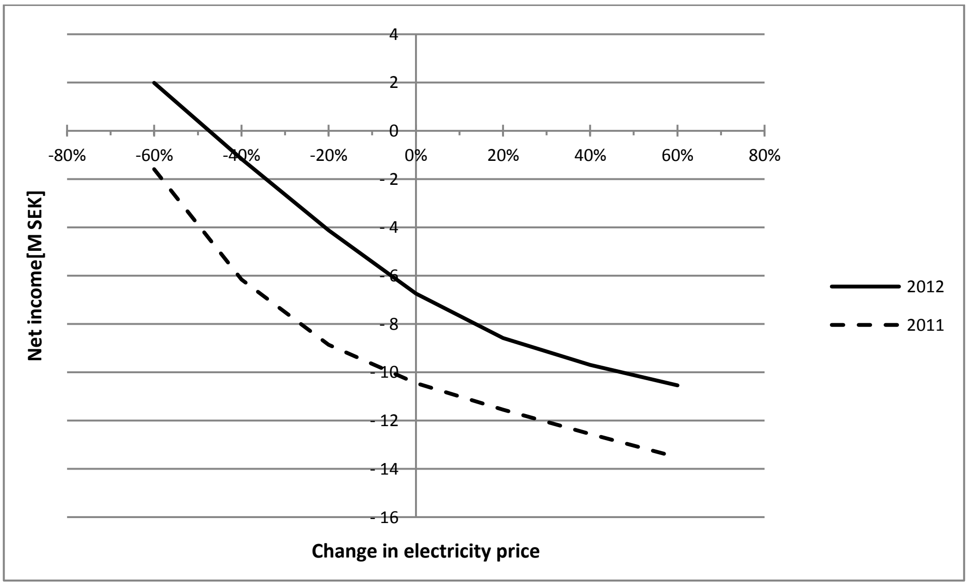


**Figure 25.** Sensitivity analysis of how changes in variable running costs affect the net income.

A 50% reduction in the electricity price would be required for 2012 to make the process profitable. For 2011 a reduction with about 70% would be required.

### *EFFICIENCY*

The efficiency of the plant has an effect on the profitability and this is shown in Figure 26.

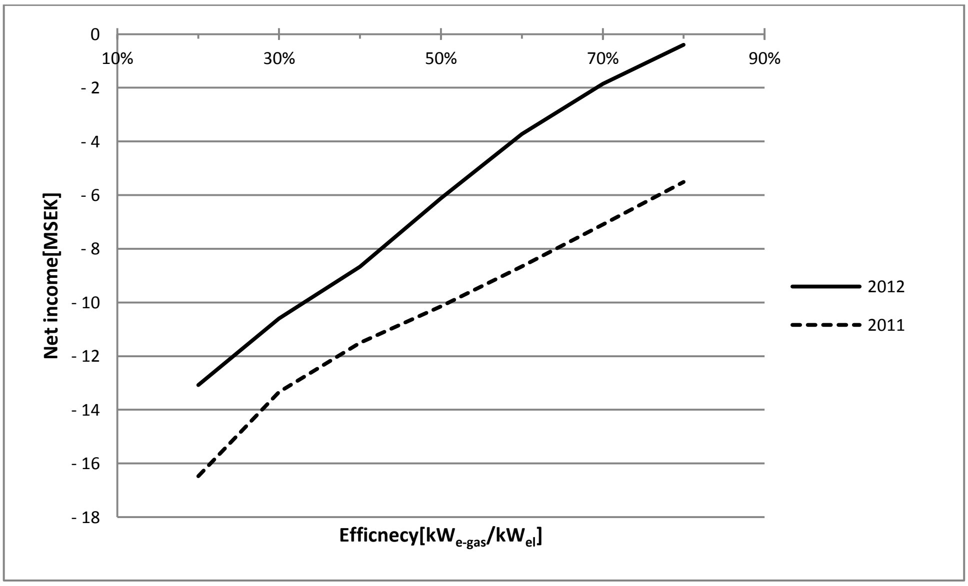


**Figure 26.** Sensitivity analysis of how changes in process efficiency affect the net income of the plant.

A higher efficiency has a positive effect on the net income, it could be reached by doing a better integration of the process of by technological development of the electrolyser. The maximum theoretical efficiency of the process, assuming the electrolyser has a maximum efficiency of 100%, is 83,1% and is the result of the inevitable heat loss in the Sabatier reaction.

## 5.2 ECONOMY OF INTEGRATING A POWER-TO-GAS PROCESS AT GOBIGAS

An evaluation of the potential of implementing a power-to-gas process in GoBiGas is presented below. The two options, $H_2$ addition in the syngas steam and implementation in separate Sabatier reactor system, are evaluated from an economic perspective and a sensitivity analysis is carried out on the most economically viable option.

### 5.2.1 INPUT

The characteristics and specifications of the input parameters are accounted for in section 3.2. This summary below deals with the details of the input variables specifically for the GoBiGas case and compares the first and second option. Table 16 shows the adjusted investment costs presented for option 1 and 2.

**Table 16.** Investment costs for the scale that is used at GoBiGas.

| Equipment | Option 1 Cost [MSEK] | Option 2 Cost [MSEK] |
|---|---|---|
| **Electrolyser** | 23,4 | 23,4 |
| **Sabatier reactor (average)** | N/A | 10 |
| **Dryer** | N/A | N/A |
| **E-gas compressor** | N/A | 2,9 |
| **Total cost** | 23,4 | 36,3 |

The total investment cost is estimated to 23,4 MSEK for option 1 and 36,3 MSEK for option 2. The input data for the economic analysis of integrating a power-to-gas process at GoBiGas is presented in Table 17.

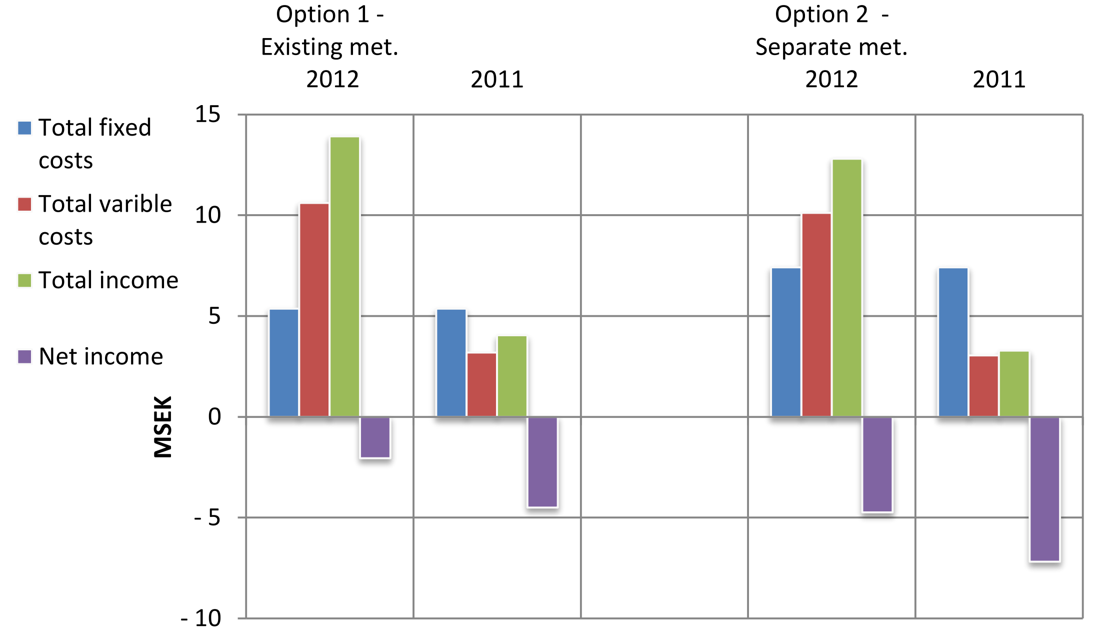
**Table 17.** Input data to the economical calculations. A more detailed description of how the different inputs were obtained can be found in Appendix A - Input Data.

| Input, costs and prices | | Option 1 | Option 2 |
|---|---|---|---|
| **Investment cost** | SEK | 23 400 000 | 36 300 000 |
| **Interest** | % | 7 | 7 |
| **Economical life time** | Years | 15 | 15 |
| **Electricity price** | $SEK/kWh_{el}$ | Hourly prices 2011 & 2012 | |
| **Electricity distribution, fixed** | SEK/year | 1 640 000 | 1 640 000 |
| **Electricity distribution, variable** | $SEK/kWh_{el}$ | 0,031 | 0,031 |
| **Gas distribution** | $SEK/kWh_{SNG}$ | 0,1 | 0,1 |
| **Operation & Maintenance** | SEK/year | 1 170 000 | 1 670 000 |
| **Water fixed** | SEK/year | 4600 | 4600 |
| **Water variable** | $SEK/m^3_{H2O}$ | 4,8 | 4,8 |
| **Price SNG** | $SEK/kWh_{SNG}$ | 0,7 | 0,7 |
| **Price $O_2$** | $SEK/kg_{O2}$ | 0,325 | 0,325 |
| **Price district heating** | $SEK/kWh_{heat}$ | 0,2 | 0,2 |

The process is run at full capacity when the running costs are lower than the income from selling the methane and the oxygen. When the running costs are higher than the income the process is shut down for option 1 and run at 20% for option 2. The electricity price used changes each hour and result in a different net income each hour of production. For the power-to-gas process at the GoBiGas plant the income comes from selling the produced e-gas, selling heat to the district-heating network and by selling oxygen that is being produced in the electrolyser. The variable costs for operation at 20% during periods with high electricity prices only applies for option 2. Due to the strategic position of the GoBiGas plant near a large water reservoir, cooling capabilities are considered abundant and cheap and are therefore neglected as variable costs. Also the deactivation of the catalyst and the subsequent variable cost of replacing are considered negligible.

### 5.2.2 RESULTS

The economic evaluation clearly shows that despite a high level of utilization 2012, the marginal profit of running the plant does not cover the high initial investment costs. As a result of this, the annual profit is negative and the process is unable to run economically with the input factors stated above. The final results of the two options for 2012 and 2012 are summarized in Figure 27.



| | | | | | |
|---|---|---|---|---|---|
| Operation[hr/year]: | 7 499 | 2 206 | | 7 159 | 1 835 |
| Annual income[MSEK]: | -2,0 | -4,5 | | -4,7 | -7,2 |
| e-gas prod [GWh/yr]: | 16,5 | 5 | | 16,5 | 7,1 |

**Figure 27.** Overview of the economics for option 1 and 2 when running process with electricity prices from 2011 and 2012. The number of operating hours and the annual income for 2011 and 2012 is shown below the chart.

The results clearly show that option 1 is the most economically efficient option, but fails to produce a profit for the input values stated. The main difference between option 1 and option 2 is the extra investment cost of the methanation process, which increases the investment cost by 43% and contributes to a higher annual fixed cost for option 2. Option 2 has a higher annual e-gas production than option 1 for 2011, this is because it is operated at 20 % during periods with high electricity prices. During these hours option 1 is completely shut down and therefore has a lower production.

Figure 28 shows the distribution of income for option 1 and 2.

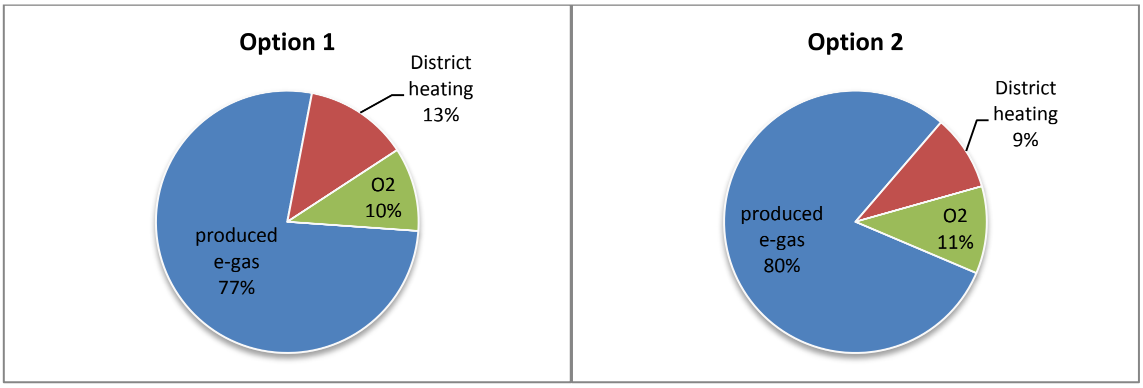


**Figure 28.** Distribution of income of option 1 and 2 for 2012.

The figure clearly shows that the main source of income for the process is from selling SNG, where the income for $O_2$ and district heating combined accounts for a total of 20 – 23%.

Figure 29 shows the distribution of costs for option 1 and 2 in 2012.

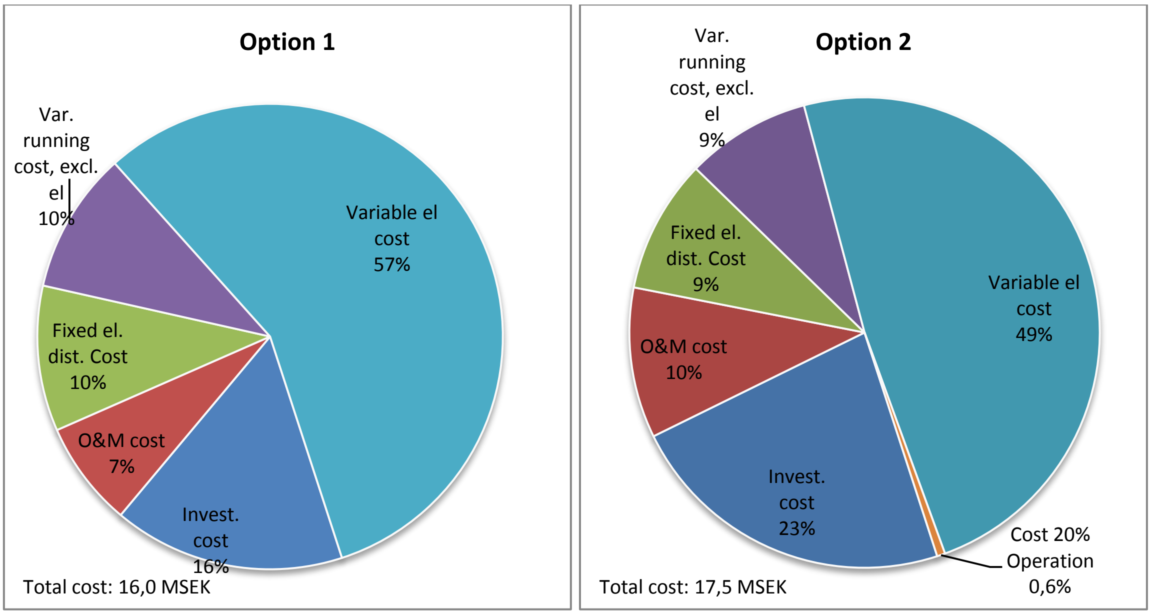


**Figure 29.** Distribution of costs of option 1 and 2 for 2012.

The figure shows that the largest part of the costs for the plant is the variable costs where the electricity cost for the electrolyser accounts for the largest share. The main difference between option 1 and option 2 is that the investment costs have a larger share in the total costs in option 2, as a result of the higher initial investment cost. Figure 29 also shows that despite the need to operate option 2 at 20% when it is not profitable to run at 100% due to technical limitation of the methanation reactor, the cost of this aspect is small compared to the total costs, accounting for only 0,6%.

### 5.2.3 SENSITIVITY ANALYSIS

A sensitivity analysis is carried out to investigate how changes in input factors will affect the final economic result of the project. The focus will be placed on option 1 as this is the option that is the closest to becoming financially viable. This evaluation is done to test the robustness of the economic evaluation but also to investigate which factors may allow the project to become financially viable in the future. The aim of the sensitivity analysis is to confirm which realistic changes in input variables are required for the project to break even.

### *INVESTMENT COSTS*

A change in investment cost will alter the yearly fixed costs in relation to the CRF. The change will not alter the variable cost and hence will have no effect on the number of operation hours, thus rendering a linear relation as shown in Figure 30.

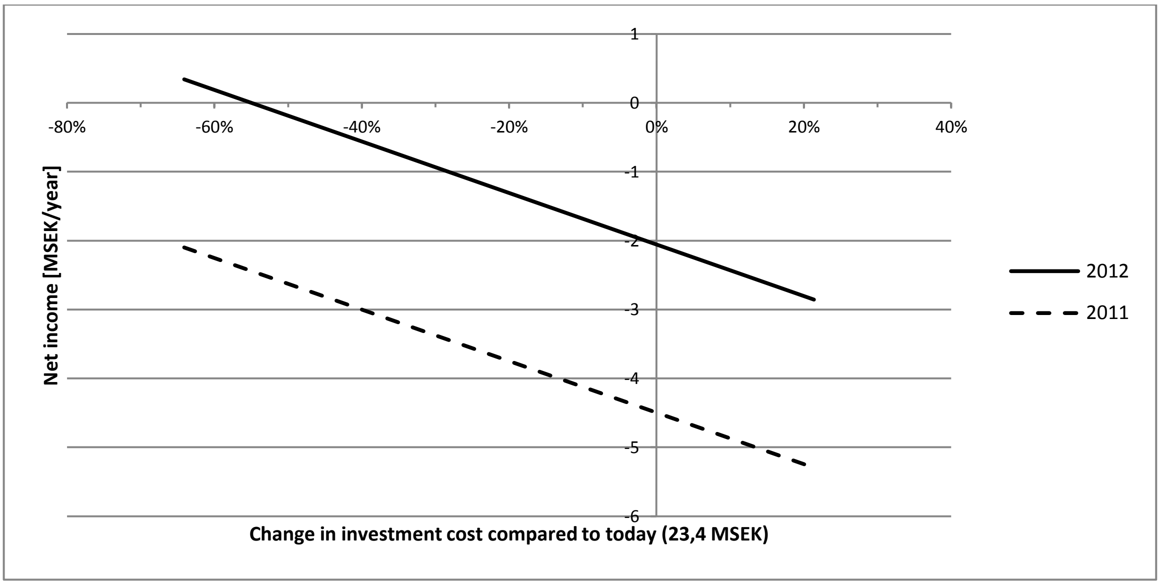


**Figure 30.** Sensitivity analysis of how changes in investment cost affect the net profit of the process.

The figure shows that a 55% decrease in investment cost from 23,4 MSEK to 12,9 MSEK would yield a break-even investment with the electricity price levels of 2012 and yield a net annual loss of 2,4 MSEK with the electricity price levels of 2011. Since the electrolyser represents the main share of the investment cost, it is clear that the economy of the whole system is strongly dependent on the price of this component, which can vary substantially depending on the supplier.

### *SNG PRICES*

A change in SNG prices will change the variable revenue of the primary product of the process, and will hence affect the number of operation hours. The number of operation hours affects the annual income and hence the relationship will not be linear, which is shown in Figure 31.

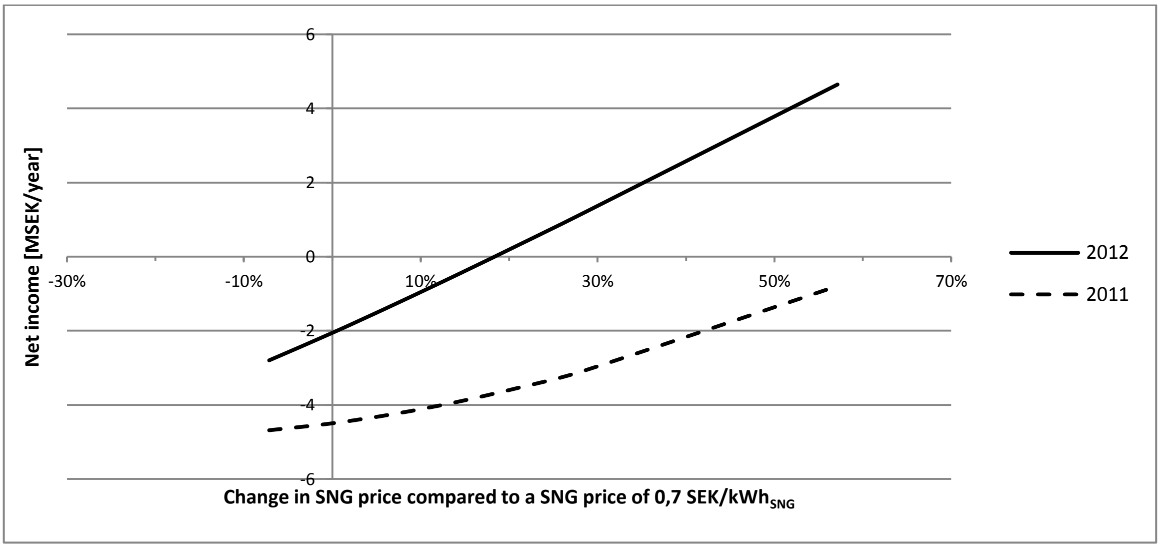


**Figure 31** Sensitivity analysis of how changes in SNG price affect the net income of the process.

The figure shows that an increase 18% in SNG price (approximately 0,83 SEK/kWh$_{SNG}$) is required for the process to become financially viable with the electricity price level of 2012. An SNG price increase of more than 55% is required for the process to become financially viable with the electricity price level of 2011. The net income curves have different gradients due to the yearly difference in hourly electricity prices.

### *ELECTRICITY PRICE*

A change in electricity price will have the effect of increasing (or decreasing) variable costs by an amount proportionate to the hourly spot price of the electricity price level of 2012. This will affect the entire price curve and will hence affect the number of operation hours. The number of operation hours affects the annual income and hence the relationship will not be linear, which is seen in Figure 32. The decrease in electricity price can be compared to an average electricity price of 0,28 SEK/kWh in 2012 and 0,43 SEK/kWh in 2011.

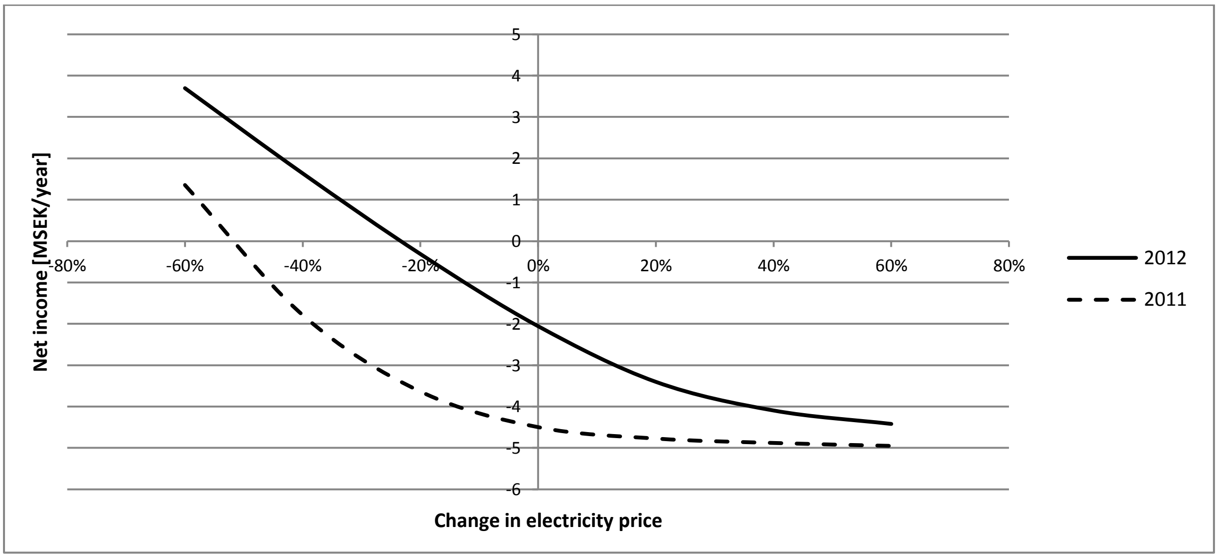


**Figure 32.** Sensitivity analysis of how changes in variable running costs affect the net income.

The figure clearly shows that a decrease in electricity price will increase the number of operation hours as well as increase the profitability of the process. A reduction in electricity price of 23% is required for the process to become financially viable with the electricity price level of 2012. A reduction of more than 50% is required with the electricity price level of 2011.

### *EFFICIENCY*

A change in the electricity to e-gas efficiency was also investigated to analyse its effect on net annual income. The electricity to e-gas efficiency is the ratio between the electrical energy added to the system and the lower heating value of the produced e-gas and depends on both the efficiency of the electrolyser as well as the conversion efficiency of the methanation process. An increase in efficiency will decrease the total running costs, and will hence affect the number of operation hours. The number of operation hours affects the annual income and hence the relationship will not be linear, which is shown in Figure 33.

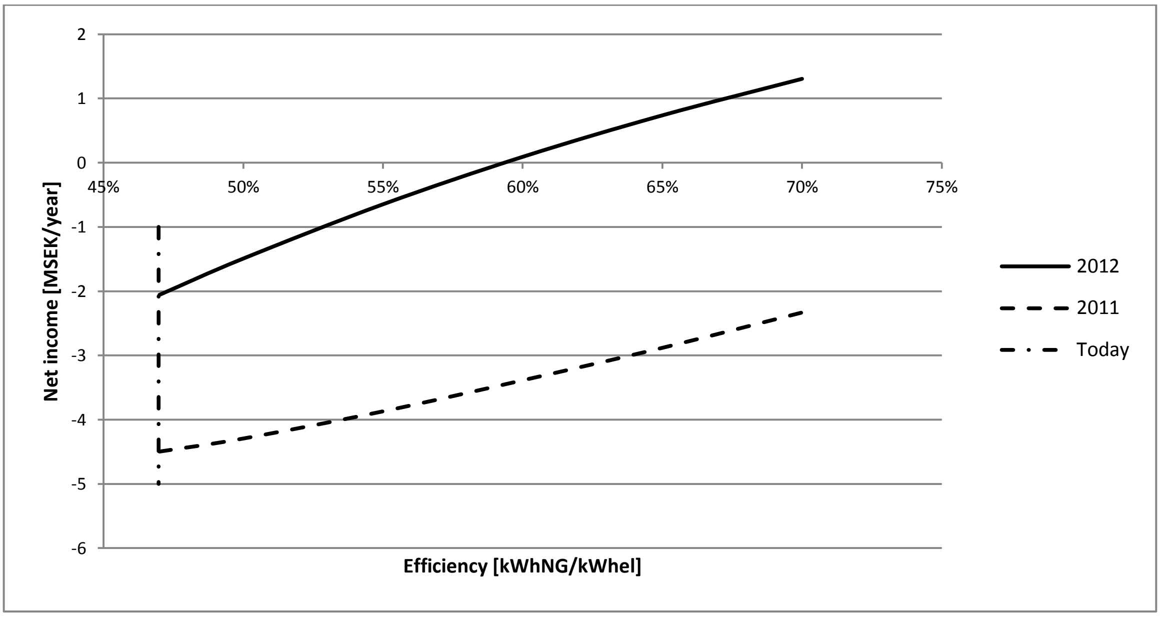


**Figure 33.** Sensitivity analysis of how changes in efficiency affect the net income.

The figure shows that an increase in efficiency will increase the net annual income as a result of an increase in operation hours as well as an increase in profit during production. An efficiency of 60% is required for the process to become financially viable with the electricity price levels of 2012 while an efficiency of more than 70% is required with the electricity price levels of 2011.

### *CONCLUSION*

The analysis shows that an increase in efficiency and e-gas price will lead to an increase in net annual income while a decrease in investment cost and fixed variable cost will also increase the net annual income. It is also clear from the analysis that the electricity price level profile has a large impact on the overall result of the production and that the change in variables carried out in this analysis are not sufficient to compensate for the higher price level of 2011. A deeper analysis of the results will be made in the discussion of the next section.

# 6 DISCUSSION AND CONCLUSION

In this discussion, Gasendal design 1(atm.) is compared with GoBiGas option 1 (existing met.). In this section the power-to-gas process at respective plant might be referred to as only GoBiGas or Gasendal.

## 6.1 TECHNICAL COMPARISON

The technical performance data is presented in Table 18 to give an overview of the two cases.

**Table 18.** Summary of the performance of power-to-gas process integrated at GoBiGas and Gasendal.

| Performance | GoBiGas | Gasendal |
|---|---|---|
| **Produced e-gas [Nm$^3$/hr]** | 220 | 340 |
| **Produced e-gas [kW]** | 2050 | 3170 |
| **Efficiency [kW$_{e\text{-}gas}$/ kW$_{EL}$]** | 47,0% | 47,7% |
| **Utilized heat [kW$_{heat\ usefull}$]** | 1190 | 705 |
| **Total efficiency [kW$_{egas+heat\ usefull}$ / kW$_{EL}$]** | 74,3% | 58,3% |

The power-to-gas process at Gasendal has about 50 % larger production than GoBiGas, because more $CO_2$ is available at Gasendal. The efficiency of the process at Gasendal is slightly higher, due to lower compression work. The total efficiency is much higher for GoBiGas, because the methanation takes place at a higher pressure and because low quality heat can be utilized in district heating. On the other hand the heat is slightly more valuable at Gasendal, where it replaces natural gas.

The degree of integration is higher at GoBiGas, where the whole Sabatier reactor system is integrated in the already existing system. This has many benefits compared to implementing the process at Gasendal, where a Sabatier reactor system is needed. The main benefit is that existing equipment can be used, including compressor, reactors heat exchangers etc. which lowers the investment cost. Another benefit is that there is no problem with shutting down the power-to-gas process completely, because the methanation reactors will still be running even if there is no addition of $H_2$ from the electrolyser.

## 6.2 ECONOMICAL COMPARISON

In Table 19 some key economical parameters are presented for the power-to-gas process at GoBiGas and Gasendal.

**Table 19.** Summary of some economical parameters between the two cases.

| Input, costs and prices | | GoBiGas | Gasendal |
|---|---|---|---|
| **Specific yearly fixed cost** | SEK/year/kW$_{e\text{-}gas}$ | 2630 | 3320 |
| **Share of income from e-gas** | [ ] | 77% | 82% |
| **Breakeven SNG price 2012** | SEK/kWh$_{SNG}$ | 0,83 | 0,97 |

The specific fixed cost (fixed cost per kW$_{SNG}$) is higher for the process at Gasendal, about 40% higher. At GoBiGas the Sabatier reactor system already exists, as well as a compressor, which lead to a lower investment cost and a lower investment also leads to a lower operation and maintenance cost.

The share of the income comes from the produced e-gas has an effect on the overall economy, the smaller share it has the better for the overall economy. It indicates how well the secondary products are utilized. Both have about 80 % of the income coming from the produced e-gas, with a slightly advantage for GoBiGas, because more heat can be utilized. On the other hand if the natural gas price would be the same as the SNG price then the picture would be very different.

The breakeven price of SNG for 2012 is lower for GoBiGas than Gasendal. GoBiGas becomes profitable at about 18% higher price than today and Gasendal at about 39% higher, this can be seen graphically in Figure 34.

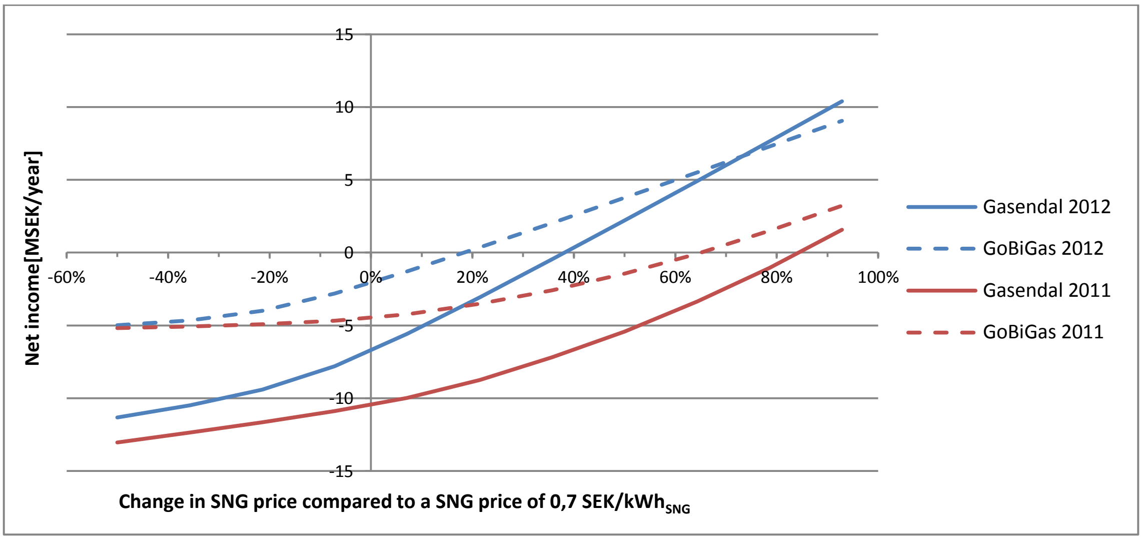


**Figure 34.** Sensitivity analysis of how changes in SNG price affect the net income, comparing Gasendal design 1 (atm.) and GoBiGas option 1 (existing met.).

Figure 34 shows that GoBiGas is the best alternative today and that it will become profitable before Gasendal. Gasendal will however become more profitable than GoBiGas if the SNG price becomes even higher because of the larger production at Gasendal.

From this analysis it is clear that integrating the process at GoBiGas would be the first alternative to consider investing in. It should be noted changes in the natural gas price and district heating price could change the picture. As an example: if the NG price would be as high as the SNG price then Gasendal would become the better choice.

# 7 FUTURE SCENARIO

A final discussion will be presented to discuss the potential of e-gas and the power-to-gas process in the future, and the factors that affect its success.

## 7.1 MODEL 2022

To evaluate the potential of implementing the power-to-gas process in a future energy market scenario, a price curve prediction for 2022 provided by Joel Goop and Lisa Göransson was used. The model treats the European energy market and produces the marginal electricity production cost with a 3-hour resolution. It is assumed that the marginal electricity production cost is a sufficiently accurate representation of the regional electricity cost. The model is based on the assumption that the national renewable energy action plans (considered “legally binding” 2020 targets for share of renewable energy use in final energy consumption.) of the EU member states are conducted as planned by 2020. The underlying method is based on linear programming with several input parameters, including an inventory database for the current European power production fleet and its lifetime (Kjärstad and Johnsson, 2007), an investment model for the implementation of different renewable energy production types (Odenberger, 2009), as well as projections of future fuel prices. The model also takes into account transmission limitations between regions, which are reflected in the resulting price fluctuations. The two cases that will be evaluated in the 2022 price model are GoBiGas – option 1 and Gasendal – design 2. All factors except the electricity price are assumed to be the same as 2012, and the price curve of 2022 can be found in Appendix A - Input Data. The results are compared to the price levels of 2012 and presented in Table 20.

**Table 20.** Comparison of economic results.

| Case / Year | GoBiGas 2022 | GoBiGas 2012 | Gasendal 2022 | Gasendal 2012 |
|---|---|---|---|---|
| **Operation Time [hrs]** | 2 609 | 7 499 | 1 962 | 6 977 |
| **Annual Income [MSEK/yr]** | -4,0 | -2,0 | -10,0 | -6,7 |

The results show that the operation time and annual income decrease for the electricity price model of 2022 compared to the electricity price level of 2012. It is also clear that the operation time decreases with 65% in the GoBiGas case and only 72% in the Gasendal case, due to the increase in average price level.

One factor that effects the outcome of the analysis, which is important to recognize, is that despite the fact that the 2022 model takes into account transmission limitations in the region, the model is an optimization, and therefore the scenario should be considered optimistic. When comparing the price curve of 2012 with the projected price curve of 2022, one can see that the price has a higher rate of fluctuation in 2022, but higher amplitude in the fluctuation in 2012. This is in part due to the optimization of the energy system made in the model and has a negative impact on the profitability of a technology that requires low prices, such as the power-to-gas process being investigated in this case. Nonetheless, the simulation gives an idea of how the price scenario may look in the future with a higher penetration of renewable production sources, and the projected higher price curve alone has a negative impact on the profitability of the power-to-gas implementation.

## 7.2 INTERMITTENT POWER GENERATION AND ITS RELATION TO POWER-TO-GAS

One argument for the power-to-gas concept is to use such plants to balance an electricity system with a large amount of intermittent energy sources. This balancing issue could be seen from different perspectives; the whole system and from the perspective of the owner of an intermittent power plant.

A system with large amount of intermittent power will have periods with much power production and periods with little production from these sources. There are many ways discussed to solve this problematic situation and one way discussed is to use power-to-gas. During periods with much intermittent power production the electricity could be used for e-gas production and periods with little intermittent power the e-gas could be used for electricity generation. This is most relevant for a future with much intermittent power in the system. It is more relevant for regions like Denmark and Germany with little regulating power, compared with Sweden, with lot of hydro power that could be used to compensate for fluctuations in the power production.

The other perspective, that of the power plant owner who is selling the electricity to the market, could be more interesting when looking at Sweden. Most electricity in Sweden today is sold and bought on the day-a-head market, elspot. The sellers and buyers tell the market how much they are going to produce and consume the next day. One difficulty with intermittent power is to predict how much electricity the power plant will produce. If the seller of electricity produces more or less than they have promised on the day-a-head market they can sell or buy the difference on the intraday market, elbas. If they fail to sell it on this intraday market they will be forced to buy regulating power to keep the electrical system in balance, which means less profit. This intraday market could be of interest for a power-to-gas plant. If the power-to-gas plant bought electricity on this market instead of the day-a-head market it could argued that it helps to balance the system and using overproduction from for example wind power to produce e-gas.

Seen from the perspective of Gasendal and GoBiGas, this intraday market could have the potential to make the process more economical. Days when it is too expensive to run the power-to-gas plant on electricity offered at the day-a-head market, the plant could buy electricity on the intra-day-market instead. The electricity on the intraday market could be lower because the sellers want to avoid having to buy regulating power, but that require that there is an overproduction. On a day when lot of power instead is requested then the price could be higher than the day-a-head price. It is therefore not wise to only buy electricity from this market but days when the plant isn’t running it would be worth considering.

## 7.3 POTENTIAL

The potential for SNG depends on several different factors, including what technology is used for production, what economic incentives and policy instruments exist and what possible applications exist. A short discussion on these subjects is presented below

The bio-gas upgrading plant at Gasendal, biomass gasification in GoBiGas and the implementation of Sabatier reactors all present different techniques for producing SNG. The technologies have varying degrees of economic potential, but it is important to realize that technologies general become cheaper over time and as bio-gas production expands and the technology becomes more efficient, the potential of SNG production will increase. For example, Lukas Grond, DNV KEMA, expects the investment cost of small to medium scale methanation solutions (<10 MW such as the one used in the investigated Sabatier Processes) to decrease to half the price of the accompanying electrolyser over the next 10-15 years as a result of commercialization and technological advances.

Another major influence is the government-stated research and development focus. A clear example of this is the “fossil free transportation” investigation currently being conducted by SOU in Sweden, whose goal is to formulate a plan for decreasing the net $CO_2$ emissions of the Swedish transport sector to 0 by 2050. Another example is legislations limiting the emissions of maritime transportation, such as the Sulphur Emission Control Areas classification of the North and Baltic sea (WSP 2013). As a result, LNG (liquefied natural gas) has received greater interest as a potential fuel for maritime transportation and LBG (liquefied bio-gas) could potentially become a non-fossil alternative.

Concerning the possible applications of SNG, it is important to take into account the alternative technologies that may compete with the expansion of bio-gas. If one looks at the application of SNG as a vehicle fuel, it is interesting to compare the wind to wheel efficiencies of competing sustainable transportation technologies. Taking into account the steps involved in converting electrical power to “drivetrain power” in a vehicle, the e-gas production has a total efficiency of 17,2% as compared with an electrical vehicle, 72% and a fuel cell vehicle, 35% (for complete calculations please see Appendix A - Input Data Efficiency Calculation). The transmission losses from production source and drivetrain losses are considered to be the same for the 3 cases and are therefore not included. With these quick calculations in mind it is clear that bio-gas is not the most efficient solution when converting electrical power to delivered power from a car engine, but a few other factors need to be taken into account when evaluating bio-gas as a fuel for transportation. First of all, bio-gas can be combusted in a regular internal combustion engine with only minor modifications, and commercial engines have existed on the market for at least a decade (WSP 2013) which is not the case for fuel cell vehicles. Also, bio-gas can be stored in a tank and can be refuelled almost instantaneously which is not the case for electric vehicles on the market today. Finally, and most importantly, bio-gas vehicles are economically viable with the technology available today, which is the case for neither fuel cell nor electrical vehicles.

SNG as an energy source of transportation also needs to compete with the existing alternatives for transportation fuel. The “at the pump” price of SNG as vehicle fuel was 1,1 SEK/kWh (WSP 2013) compared to gasoline and diesel, at 1,60 and 1,43 SEK/kWh (SPBI) respectively in late 2012, which also shows the potential of bio-gas as a major transportation fuel. It is important to notice, however, that bio-gas is currently exempted from energy tax and the efficiency of today’s gasoline and diesel engines are slightly higher.

Other applications for bio-gas exist as well, including CHP, district heating production and other industrial applications where natural gas can be replaced. In these applications however, the taxes on the natural gas is lower than for vehicle fuel and the bio-gas solution becomes relatively more expensive. Nonetheless, bio-gas can in practice replace natural gas in most applications and with the right policies and economic incentives the possibility definitely exists.

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# APPENDIX A - INPUT DATA

## ELECTROLYSER

Here follows a technical description of the commercially available electrolyser used in the calculations: NEL A - atmospheric electrolyser. This electrolyser comes from the manufacturer NEL hydrogen, a Norwegian company with long experience of electrolysers. The properties of this electrolyser are presented in Table A1. The power consumption includes all auxiliary equipment except high pressure compression. The electricity consumption decreases linearly down to 20% of the max capacity; this is because of lower current densities (Harrison and Levene 2008). The cost of this electrolyser is about 12,7 MSEK for a capacity of 500 $Nm^3_{H2}$/hr. This electrolyser can switch between operations at 100% to 20% in about 10 minutes. A cold start to 100% takes several hours (Taalesen 2013). Table A1 shows a summary of the properties of the electrolyser.

**Table A1.** Properties of the NEL A – Atmospheric electrolyser.

| **NEL A - ATMOSPHERIC ELECTROLYSER** | |
|---|---|
| **Capacity/Nominal Flow Rate Capacity range ($Nm^3$ $_{H2}$/hr)** | 10 - 500 |
| **Power consumption at maximum capacity [a] (kWh/$Nm^3$ $_{H2}$)** | 4.833 |
| **Operation, % of max capacity** | Automatic 20 - 100% |
| **Power consumption at 20% of maximum capacity [a] (kWh/$Nm^3$ $_{H2}$)** | 4,333 |
| **Operating temperature** | 80°C |
| **$H_2$ purity (%)** | 99.9 ± 0.1 |
| **$O_2$ purity (%)** | 99.5 ± 0.1 |
| **Electrolyte** | 25% KOH aqueous solution |
| **Feed water consumption** | 0.9 litre / $Nm^3$ $H_2$ |
| **Price, 500 $Nm^3$ $_{H2}$/hr (MSEK)** | 12,7 |
| [a] *Includes auxiliary equipment, except high pressure compressor* | |

Another manufacturer is a Chinese company called Zhengzhou Yukun Machinery Equipment Co., Ltd. Their electrolyser is able to start from being switched off more rapidly than the NEL, other than that it has slightly lower performance than the NEL electrolyser but a lower price. The price is 7,8 MSEK for a size of 500 $Nm^3_{H2}$/hr (Jiang 2013).

## ECONOMY

This section presents the different costs and the prices of the products relevant to a power-to-gas plant. They are used as input data to the economical analysis of integrating power-to-gas at Gasendal and GoBiGas.

### COSTS

#### *CAPITAL/INVESTMENT COST*

The major components in a power-to-gas plant are: an electrolyser, a Sabatier reactor system, a dryer and a SNG compressor. In Table A2 is the investment cost presented for a specific size of each component. Note that the sizes need to be adjusted before they are added to the total cost; this is done according to equation (8) in the method.

**Table A2.** Investment costs of equipment used in the SNG process [a] (Taalesen 2013), [b] (Culmsee 2013), (Grond 2013), [c, d] (Strandberg 2013).

| Equipment | Size | Cost [MSEK] |
|---|---|---|
| **Electrolyser** [a] | 500 $Nm^3H_2/hr$ | 12,7 |
| **Sabatier reactor** [b] | 375 $Nm^3_{SNG}/hr$ | 9 – 19 |
| **Dryer** [c] | ≈1000 $Nm^3_{CH4}/hr$ | 0,4 |
| **SNG compressor** [d] | ≈500 $Nm^3_{CH4}/hr$ | 1 |

### ELECTRICITY PRICE

Nord pool provides historical data for the electricity price each hour of the day. The price curves for Sweden price area SE3 for the years 2011 and 2012 can be seen in Figure A1 (Nord Pool Spot 2013). The electricity used for this type of process is tax free (Swedish Law on Energy Taxation 1994).

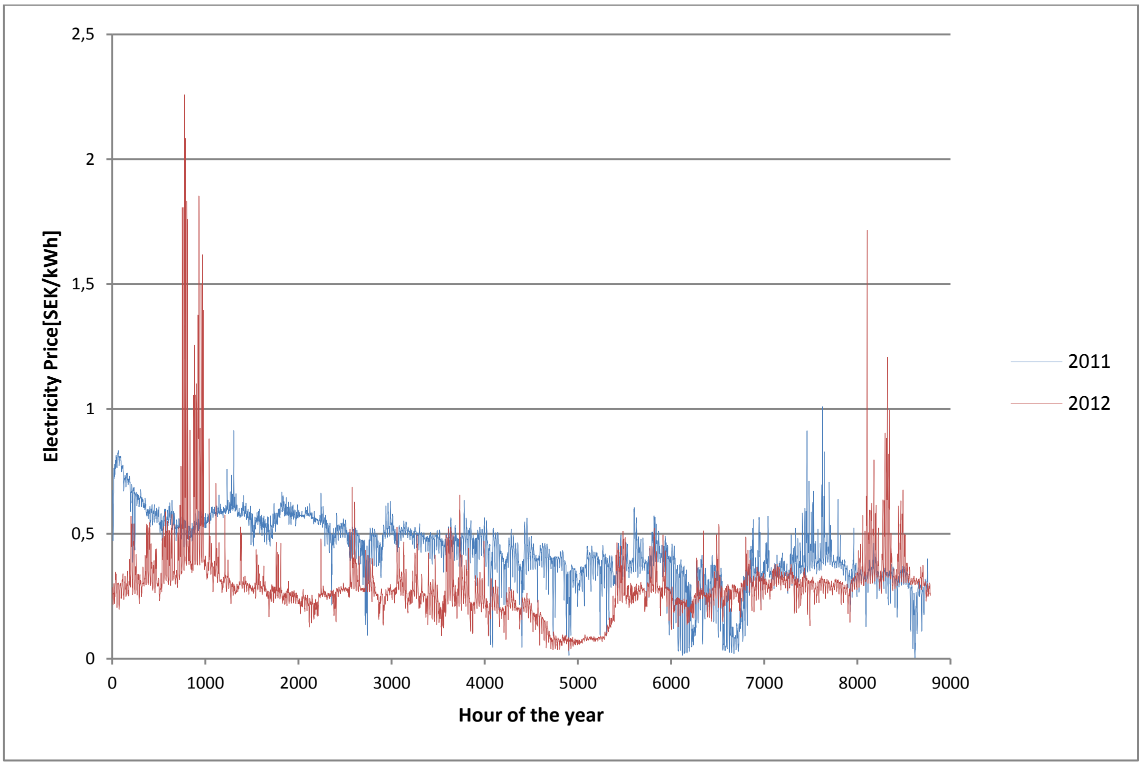

**Figure A1.** Electricity price each hour of the year for 2011 and 2012.

### ELECTRICITY DISTRIBUTION COST

There is a cost to have the electricity distributed to the site where it is used, the electricity distribution cost. The cost of delivering electricity within Göteborg Energi's network is presented in Table A3. The power-to-gas process will be using the subscription for High voltage 10 kV. The electrical boiler is presented for comparison but is no longer available.

**Table A3.** The different types of subscriptions for electricity distribution available at Göteborg Energi and the costs associated with each subscription (Göteborg Energi AB 2012).

| Power subscription | Subscription SEK/year | Power transfer SEK/kWh | Power SEK/kW, month | Reactive Power SEK/kVAr, month |
|---|---|---|---|---|
| **High voltage 10 kV** | 8 800 | 0,031 | 31,6 | 7,0 |
| **Electric boiler 10kV*** | 18 800 | 0,048 | 0 | 0 |
| ** able to shut down when there are high demands on the system, this subscription is no longer available, this information is from 2011.* | | | | |

### *Other Costs*

Other costs relevant for a power-to-gas plant are presented in Table A4, including: natural gas distribution, water, operation and maintenance.

**Table A4.** Other costs relevant for a power-to-gas plant [a] (Hedgran 2013), [b] (Göteborgs stad 2012).

| Cost | Unit | |
|---|---|---|
| **Natural gas distribution [a]** | SEK/kWh$_{SNG}$ | 0,1 |
| **Water consumption, fixed [b]** | SEK/year | 4621 |
| **Water consumption, variable [b]** | SEK/m$^3$ | 4,8 |
| **Operation & maintenance** | % of investment | 5 |

### *Income*

The power-to-gas process will generate income mainly from the produced methane but also from other products like oxygen and district heating. In Table A5 are the values for these products presented.

**Table A5.** Income for a power-to-gas plant [a] (Zinn 2013), [b] (Saxe and Alvfors 2007), [c] (Mohseni 2012).

| Income | Unit | |
|---|---|---|
| **SNG [a]** | SEK/kWh | 0,7 |
| **Saved NG [a]** | SEK/kWh | 0,25 |
| **Oxygen [b]** | SEK/kg | 0,325 |
| **District heating [c]** | SEK/kWh | 0,2 |

### *SNG and NG Price*

The SNG and NG prices are based on the lower heating value. The SNG price is based on a future estimation of the price of bio-gas. The NG price is based on the expected NG price in the near future (Zinn 2013).

### *Oxygen*

The electrolyser will produce a stream of oxygen that could be of interest to capture and sell to an external buyer, to increase the income from the plant. This will require extra equipment and higher electricity consumption therefore some additional costs. The extra costs associated with the oxygen process is accounted for in the price of the oxygen, to make it easier to monitor this factor and also because the cost is quite uncertain. The cost of this oxygen process equipment will depend on how the oxygen is delivered to the buyer. If high compression is required and long transportation, then the cost for the process will be higher than if only a small pressure increase is required. A smaller pressure increase would be likely if the oxygen is delivered to a close by industry in pipes instead of tanks. The most common users of oxygen are metal processing industries (65% of the oxygen use), pulp and paper industries (15%) and chemical industries like petroleum industries (15%) (Saxe and Alvfors 2007). Both Gasendal and GoBiGas are located relatively close to petroleum industries, there is therefore a potential that there might be a close by buyer.

The value of the oxygen is assumed to be 0,51SEK/$kg_{O2}$ (Saxe and Alvfors 2007) this value is based on the total delivery income from oxygen and the total oxygen delivered in 2003 in Sweden.

The costs associated with handling the oxygen are estimated and used to adjust the oxygen price. As a base case for the economic analysis the value of the oxygen after all costs associated with the oxygen are accounted for, is estimated to be between 0,30-0,35 SEK/$kg_{O2}$. To reach this number it is assumed that the plant is operated 5000 hrs/year, the investment is 3 MSEK, the electricity consumption is 70 - 220kW and an electricity price of the average of 2012.

### *DISTRICT HEATING*

The possible income from district heat is difficult to estimate since the selling price of district heating is based on the instantaneous marginal cost of production in the district-heating network, much like the instantaneous electricity price is based on the marginal cost of production.

A previous study has assumed that the marginal production cost of district heating is equal to the average fuel cost for district heating in 2007, and has therefor set the value to 0,278 SEK/kWh (Mohseni 2012). Another previous study assumed the income from district heating (in this case waste heat specifically) to be 50% of the selling prices to customers (Arnell, et al. 2012) which in the Göteborg region would amount to 0,358 SEK/kWh (Svensk Fjärrvärme).

It is assumed that the district heating prices follow the electricity prices on a season basis, and that therefore the district heating prices are the lowest when the electricity prices are the lowest. Since the production of SNG will only run when electricity prices are sufficiently low, this means that the income from district heating for the excess heat will most likely be lower than the yearly average.

With this motivation, the income for district heating is assumed to be 0,20 SEK/kWh, which is approximately 30% lower than the estimate made by Mohseni. The assumption will be counted for both medium quality heat, with a temperature between 100°-300°, as well as high quality heat, with a temperature above 300°. This assumption is based on the fact that no clear consumer of intermittent high quality heat can be identified in the area and therefore the high quality heat will be used in district heating. A higher income is possible if a high quality heat consumer is identified.

## PRICE CURVE OF 2022 MODEL

The price curve for the 2022 model was taken from a report currently being written by Joel Goop and Lisa Göransson. The model consisted of three consecutive weeks for the 4 seasons and the model was therefore repeated to create a price curve for the whole year, which can be seen in Figure A2. The time resolution is 3 hours.

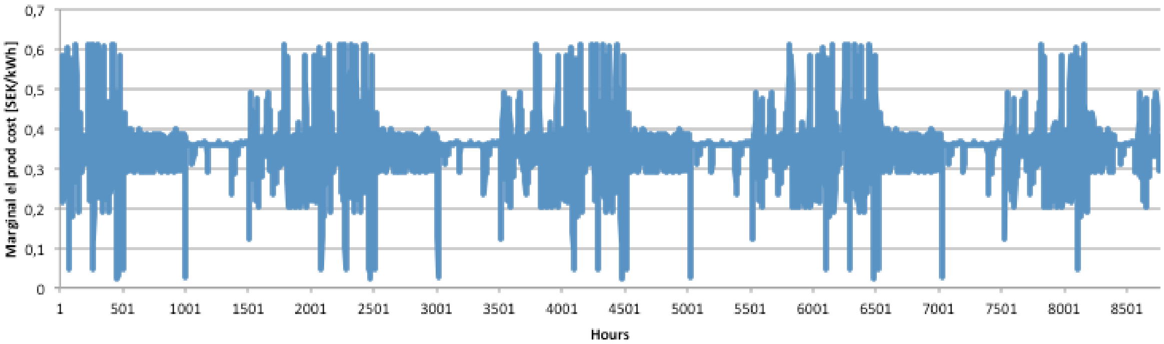

**Figure A2.** Price curve of 2022 model used in evaluation of future potential.

## $CO_2$ REMOVAL

The heat needed for the $CO_2$ removal was determined from actual operation data in the following way:

Data from Gasendal operation 2012 average(Göteborg Energi):

Power to steam production (from burning natural gas): 690 kW

Biogas received from Gryaab: 1018 Nm3/hr

Biogas composition: 2/3 $CH_4$ and 1/3 $CO_2$

$$CO_2 flow = 0{,}333 * \frac{1018}{3600} = 0{,}0942\ Nm^3/s$$

Assume ideal gas and normal conditions as 15°C and 1 atm.

$$\dot{n} = \frac{P\dot{v}}{RT} = \frac{101325 * 0{,}0933}{8{,}3145 * 288{,}15} = 3{,}98\ mol_{CO_2}/s$$

$$M_{CO_2} = 44{,}01\ g/mol$$

$$massflow, \dot{m} = \dot{n} * M_{CO_2} = 175{,}3\ g/s = 0{,}175\ kg/s$$

$$Energy\ required\ per\ kg_{CO_2} = \frac{700}{0{,}175} = 4000\ kJ/kg$$

## EFFICIENCY CALCULATION

E-gas:

Electrolyser efficiency: 70%

        $H_2$ – $CH_4$ conversion efficiency: 82%

        Internal combustion engine efficiency (gas): 30%

        Total efficiency: 17,2%

    Electrical Vehicle

        Charging efficiency: 80%

        Operation efficiency: 90%

        Total efficiency: 72%

    Fuel Cell Vehicle

        Electrolyser efficiency: 70%

        Fuel-cell operation efficiency: 50%

        Total efficiency: 35%

Transmission losses from power plant to vehicle and drivetrain losses of vehicles are assumed to be identical for the three cases.