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A feasibility study for electrofuel implementation with waste-to-energy plant

Master’s thesis in Sustainable Energy Systems

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MASTER'S THESIS

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

Electrofuel or Power-to-Gas is paid by great attentions nowadays since it could mean to reduce the fossil fuel dependency by renewably production of combustible gas. Main application of gas could be to supply a clean fuel for vehicles which means if the technology is feasible, share of greener energy in transportation sector's energy mix would be greatly increased. Despite having many studies focused to implement the technology with biogas facilities, very few concentrate on feasibility to apply the same technique with conventional combustion process like waste-to-energy process. Hence in this study, the electrofuel technology of Sabatier reaction, which uses renewable electricity, water and carbon dioxide to produce methane, is examined by integrating with a case study waste-to-energy plant.

The study mainly considers the economic feasibility of the combined process of Sabatier and carbon capturing process while perspectives on technical term are also discussed in parallel. Capacity of the process is varied with extreme waste-to-energy plant's electricity and emission outputs to define the gas production potentials. Profitability of different process sizes is determined as well as sensitivity analysis on key economic parameters such as synthetic natural gas (SNG) price, electricity price or change in investment cost. The work also takes into the account some change due to season that has significant effects to both rate of energy production and relevant profitability.

The results show that the integration is feasible especially when the SNG price is high which is likely in the near future. When the process size increases, the change in investment cost plays less important role in total process cost. It can also be observed that such change in production capacity, break-even points of SNG price decreases while reverse trends can be gained in case of electricity price. Besides variation on key economic parameter, another critical factor is electrolyzer efficiency which operating cost can be significantly lower when more efficient electrolyzer is used.

Keywords: Power-to-gas, Sabatier Reaction, Waste-to-energy, Cost-analysis

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1. Introduction

Trend of the future energy systems is now shifting towards for more sustainable ways. Rapid developments for renewable energy production is now being implemented worldwide as increasing share on wind and photovoltaic power production reduces significantly the anthropogenic greenhouse gas emissions which as a result would decelerate an increasing in global temperature. In Europe, key targets have been set to reduce 20% of greenhouse gases (GHG) emission as well as to increase share of renewable energies to be 20% in total energy consumption by 2020[1]. This could greatly stimulate use of the renewable power production throughout the region.

Despite having a bright future, the renewable power production still has many challenges to overcome. One of those challenges is intermittency, which means that the production could occasionally dominate the whole energy production. The displeased thing is that such occasion always lead to some surplus of electricity which usually end up being disposed. While in some period of time, the renewable energy gains so small amount of energy due to the weather uncertainties, the conventional power productions like coal-fired or gas boiler therefore have to turn on in order to cope with the demand. This implies that the larger share of the renewable energy production could create additional energy supply complexities due to the temporal and spatial fluctuations of the systems. To solve the problem, excess energy storage is then needed. The idea is that when the renewable power production gains its surplus, it should somehow be able to store the surplus and use it whenever the energy production is lowered and cannot anymore handle the demand. With such storage system, it could lead the renewable way of power generation to become more stable and practically applicable.

There are many alternatives to store the surplus of electricity: from using electrochemical battery to pumped hydro storage system. The latter means to use the surplus to pump water to the elevated reservoirs and make use of potential energy whenever there is a need [2]. But what is of interest nowadays is called Electrofuel or Power-to-Gas technology. The technology uses the surplus to produce hydrogen (H_2) which later can be used to produce methane (CH_4) by methanation reaction called Sabatier reaction. Both types of gas are very useful for being either chemical feedstock or fuels due to their energy-intensive properties [3], [4]. For methane, it needs carbon dioxide (CO_2) for its methanation reaction. Hence, CO_2, which is normally emitted from any of combustion process, could now be recycled and utilized for this purpose which would lead to a useful way to deal with CO_2 emission.

Alongside with the rapid renewable energy development, findings for more sustainable waste management is also paid by great attentions. One of the solutions is a process called waste-to-energy. The process recovers heat by combusting municipal solid waste (MSW) where the heat gained can both be utilized in forms of electricity and district heating. Since up to 60 percent of its feedstock consist of non-fossil-based materials [5], it could be, to some extent, defined as a more sustainable option of energy supplying.

To minimize loss of energy, most of the waste-to-energy plants are located close to municipalities. It means that if a technology like electrofuel can be introduced with the plant, then methane could be synthesized locally and maybe be able to supply all the gas requirements within the municipality (e.g. fuel for public transport or for immediate electricity demands during peak hours etc.). As a result, this could actually be seen as a decentralized way for the local's energy security as well as to be energy import-independent.

In this study, a selected pilot-size of the electrofuel process from a previous study will be initially used as a base case. By using such size, it is possible to identify specific needs and from that scale up the process to possible plant capacities for implementing the electrofuel process with the specific waste-to-energy plant that this study focuses on. In addition, the analysis also takes into account the economic aspects, to elaborate under what circumstances the integration of the process is economically feasible.

Problems description and aim of this study

This study aims to investigate feasibility of integration of electrofuel process with a case study waste-to-energy plant. The process size is selected from a study done by Mohseni et al [6] which then is scaled up in order to fit the actual operating conditions within the waste-to-energy plant. Size of the process is therefore varied by using different plant's extreme conditions (e.g. the size based on maximum salable electricity produced or maximum utilizable CO_2 emission). The cost-analysis will show if the process is feasible and also indicate what the most influential process parameters. The thesis also investigates some specific issues that may occur from integrating electrofuel with the waste-to-energy plant.

Delimitations of this study

Due to many possible way to utilize methane, this study will not include different ways of methane usage as a factor for the feasibility of an integration of electrofuel production process.

The study will neither consider what should be the source of electricity and related costs for such source. Instead, the electricity price gained from Nordic commodities market is used together with other relevant price sources to represent the overall electricity cost the process needs to take into the account. The electricity price will also be varied in a sensitivity analysis.

2. Background

Electrofuel Technology

Electrofuel technology is the technology that converts renewably-produced electricity into useful fuel (e.g. methane or methanol). The system has an electrolyzer where the electricity is applied to turn water into hydrogen and oxygen. The generated hydrogen is then sent to a reactor where it is mixed with carbon dioxide from any source (e.g. raw biogas or combustion process) to be reacted and yield a fuel compound. In this study, a methanation reaction called Sabatier reaction, is used to represent a way to produce electrofuel. In addition, the reaction is highly exothermic which means heat generated from the reaction together with heat left from the electrolyzer can potentially be utilized as district heat or preheating process [6]. More details about the Sabatier reaction are discussed later in this section. A simplified diagram of the process is shown as in Fig. 2.1

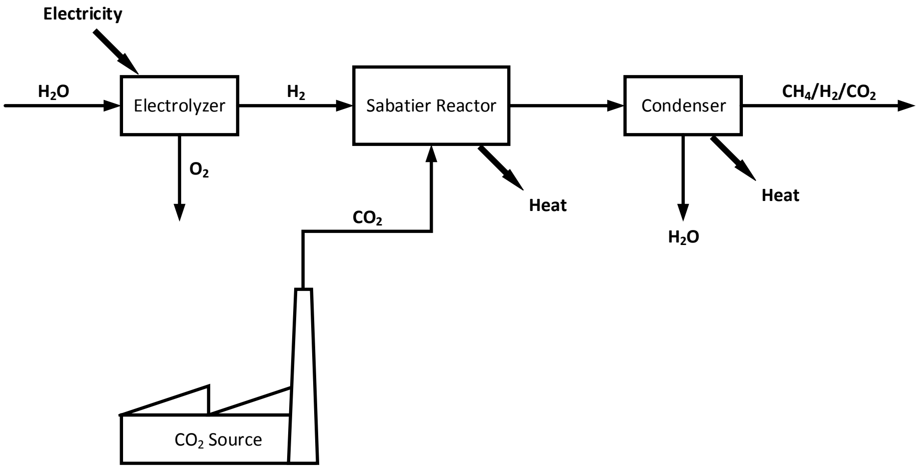


Fig. 2.1 Block diagram of electrofuel process based on lower heating value

Water electrolysis

Water electrolysis is one of the most important part of the electrofuel process. It applies electricity to split water (H_2O) into two basic compounds: hydrogen and oxygen (O_2). The overall splitting reaction is shown as Eq.(2.1).

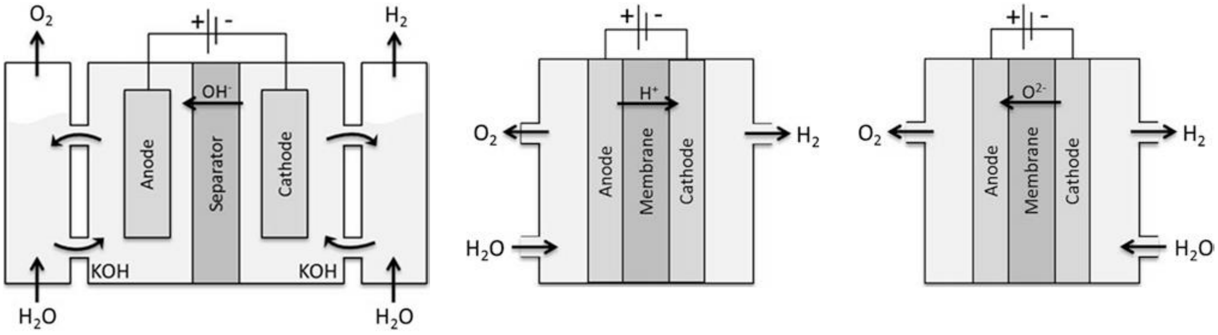
$$H_2O \rightarrow H_2 + \frac{1}{2}O_2 \qquad V_{rev} = 1.23\ V \quad (2.1)$$

V_{rev} represented in the equation is called "reversible voltage" or the minimum potential that the reaction requires for starting. However, the potential is definitely not enough for the reaction to occur, additional heat in form of higher potential, is thus needed. Hence, another voltage value of thermo-neutral voltage, V_{th}, is introduced and defined as in Eq.(2.2).

$$V_{th} = \frac{\Delta G}{nF} + \frac{T\Delta S}{nF} = 1.48\,V \tag{2.2}$$

ΔG represents term of reversible voltage while TΔS stands for the additional heat. If the considering electrolysis cells applies potential that is lower than V_{th}, then some additional heat is needed from surrounding environment for the reaction to occur. While if the same potential is equal to V_{th}, then it means that heat generated from the cells is enough for the reaction and there is no need to exchange heat with the surrounding. In the last case where the applied potential is higher than V_{th}, then some surplus heat would be gained and an appropriate cooling is required to maintain system's temperature [2].

There are three main types of electrolyzer being used nowadays: Alkaline (AEC), Polymer Electrolyte Membrane (PEMEC) and Solid Oxide Electrolyte (SOEC) [7]. The schematic diagrams of these three types of electrolyzer are shown in Fig. 2.2. In this study, only AEC is considered since it is the well-developed and commonly commercialized for any of industrial requirement [8].



(a) (b) (c)

Fig. 2.2 Different type of electrolyzers (a) AEC, (b) PEMEC, and (c) SOEC [1]

As can be seen in Fig. 2.2(a), AEC cell consists of two electrodes which are submerged in a potassium hydroxide (KOH) electrolyte with a separator that divides anodic and cathodic zones. The reactions within the cell are:

Cathode $$2H_2O + 2e^- \rightarrow H_2 + 2OH^- \tag{2.3}$$

Anode $$2OH^- \rightarrow \frac{1}{2}O_2 + H_2O + 2e^- \tag{2.4}$$

This mechanism generates hydrogen with 99.5% in purity [2], [9] with a system efficiency range between 59-80% based on higher heating value [2].

Sabatier Reaction

The Sabatier reaction was discovered in 1902 by the French chemist Paul Sabatier. The overall reaction, as can be seen in Eq.(2.5), converts CO_2 and H_2 into CH_4 and water in presence of catalyst.

$$CO_2 + 4H_2 \leftrightarrow CH_4 + 2H_2O \qquad \Delta H_R^0 = -165\ kJ/mol \qquad (2.5)$$

The reaction can be described in detail by two consecutive reactions [2] as

$$CO_2 + H_2 \leftrightarrow CO + H_2O \qquad \Delta H_R^0 = +41.2\ kJ/mol \qquad (2.6)$$

$$CO + 3H_2 \leftrightarrow CH_4 + H_2O \qquad \Delta H_R^0 = -206.2\ kJ/mol \qquad (2.7)$$

The operating temperature of the reaction is in a range of 200-400 °C. According to Lunde [9] and Brooks et al [10], the desirable operating temperature for the reaction is around 200-370 °C at the atmospheric pressure. One of the studies, however, indicates that the operating pressure can be also as high as 10 bar [10]. If the operating temperature is too high, then the reactions is reversely shifted and yields more carbon monoxide (CO) [11].

Early modern researchers aimed to implement the reaction for life-supporting purpose on spacecrafts or space stations [11]–[14]. The concept is to produce O_2 for astronauts and recirculate CO_2. Once the O_2 is consumed and CO_2 is produced from respiration, the CO_2 together with H_2 from electrolyzers are fed to the reactor to reproduce water, while generated CH_4 is disposed into the space [13]. This would create a loop which every vital substance is circulated. However, due to CH_4 rejection, mass flows within the loop could continuously decrease. Hence, in order to maintain the cycle, some carbon-based substrates and water is needed to be added to the system. In fact, this can be done easily by frozen food supply [13].

At present, the reaction is getting more attentions as a promising way to make use of CO_2 to renewably generate CH_4 and excessive heat. Therefore, integrating the Sabatier reaction into power-to-gas system would be more advantageous than having electrolyzers alone. This is because CH_4 is much easier to handle, compared to H_2, since it can use the existing gas network to transport without any further modification [2] and much lower pressure (or temperature in case of liquefaction) is required to compress CH_4 compared to H_2. In addition, volumetric heating value of CH_4 is also about 3 times higher than that of H_2 which means that with the same volumetric flow, CH_4 stream contains much higher energy than the hydrogen. Nevertheless, the main disadvantage of integrating the methanation process is conversion losses. The process' efficiency is considerably lower than just to produce H_2.

Table 2.1 Overall efficiency of all power-to-gas possibilities [2], [16]

Path	**Efficiency (%)**	**Boundary conditions**
Electricity to gas		
Electricity → Hydrogen	54-72	Including compression to 200 bar (underground storage)[2]
Electricity → Methane (SNG[1])	49-64	
Electricity → Hydrogen	57-73	Including compression to 80 bar (feed in gas grid for transportation)
Electricity → Methane (SNG)	50-64	
Electricity → Hydrogen	64-77	Without compression
Electricity → Methane (SNG)	51-65	
Electricity to gas to electricity		
Electricity → Hydrogen → Electricity	34-44[3]	Conversion to electricity: 60%, including compression to 80 bar
Electricity → Methane → Electricity	30-38[3]	
Electricity to gas to combined heat and power (CHP)		
Electricity → Hydrogen → CHP[4]	48-62	40% electricity and 45% heat, including compression to 80 bar
Electricity → Methane → CHP	43-54	

[1] SNG = Synthetic natural gas

[2] Typical pressure for compressed natural gas (CNG) used in some road-based vehicle

[3] The numbers are obtained by multiplying efficiency of Electricity → Hydrogen or Electricity → Methane (SNG) with gas conversion to electricity of 60%

[4] CHP = Combined heat and power

Catalyst

Catalysts are needed for the Sabatier reaction. A study done by Lunde and Kester (1974) reviewing that ruthenium and nickel are the most active ones for the reaction. They also mentioned that the ruthenium-based catalyst has less technical problem compared to the nickel: it has no problem with sulfur poisoning, it does not need to startup in pure hydrogen stream, and it has no coke formation problem at the high operating temperature range. Ruthenium, however, is far more expensive than nickel which makes it less attractive to use in an industrial scale [17].

Carbon capturing technology

Like other combustion processes, air is fed to the furnace within waste-to-energy process for the oxidation reaction. It means that great proportion of nitrogen (N_2) and O_2, which are discarded from the combustion process, would be in the post-combustion flue gas. Hence, to utilize flue gas as a feed for the Sabatier process, one needs to consider ways to separate CO_2 from the stream of flue gas. Same separation concepts as in biogas upgrading and coal- or natural gas-carbon capture can thus be applied for this purpose.

There is a considerable range of technologies that can be used for this purpose [18], but vacuum pressure-swing adsorption (VPSA) and monoethanolamine (MEA) absorption are most recognized and considered good alternatives.

In contrast, there is a study showing that it is possible to feed mixture of N_2 and CO_2 directly to the methanation reaction [19]. The presence of N_2 does not disturb the reaction due to very high CO_2 -selectivity of the catalyst. It is O_2 that left from the combustion chamber as a result of high air-fuel feeding ratio that is problematic. Such O_2 can react with H_2 from the electrolysis and produce water. This would decrease the overall efficiency of the system since yield of methane is lowered. Furthermore, as there are additional flows from N_2 and O_2 to the Sabatier reactor, it implies that there would be a complex separation on the post-Sabatier-process stream to divide CH_4 from all other products. If not separating away the N_2 and O_2, the volume of incoming gas flow (the feeding stream) also leads to that the process' reactor, compressor, as well as pipelines and fittings are required to be much larger in size. Hence, it can be said that the process with directly-fed flue gas is not economical in practical use and is not considered in this study.

Vacuum pressure-swing adsorption (VPSA)

VPSA uses difference in adsorption forces of different gas components to separate a specific gaseous substance from others. The feeding gas is pressurized to the desired operating pressure and adsorbed inside an adsorbent-loaded vessel. Unwanted components are adsorbed while a purified one leaves the vessel. The adsorbent is regenerated by depressurizing the vessel to the lowest operating pressure. Those unwanted components are then desorbed and purged out. In order to make a continuous gas production, the process needs at least two vessels for the operation: one for the adsorption and another one for the regeneration. By this, the two separations will occur simultaneously [20]. The basic diagram of the process can be seen as in Fig. 2.3.

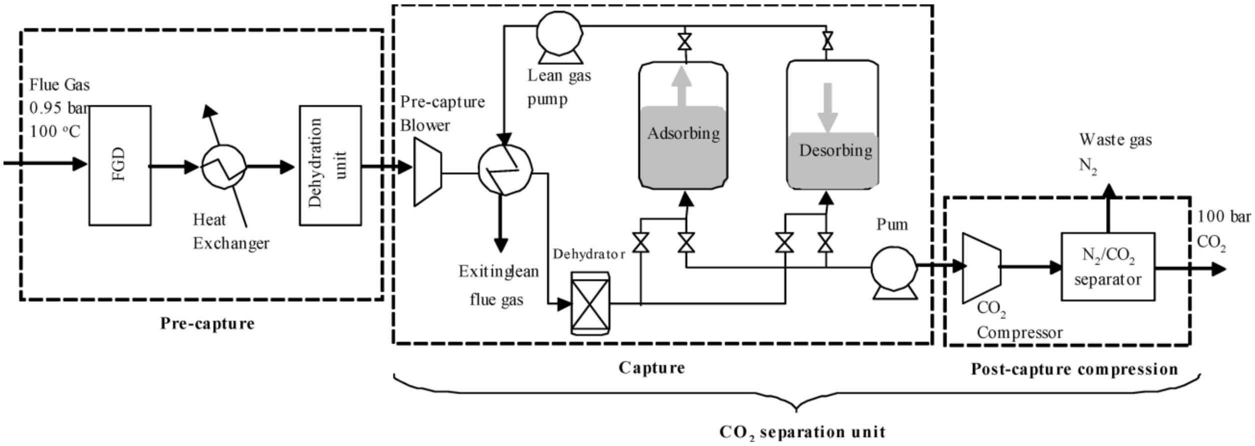

Flue Gas
0.95 bar
100 °C
FGD
Heat Exchanger
Dehydration unit
Pre-capture Blower
Lean gas pump
Adsorbing
Desorbing
Exiting lean flue gas
Dehydrator
Pum
Waste gas
N_2
100 bar
CO_2
N_2/CO_2 separator
CO_2 Compressor
Pre-capture
Capture
Post-capture compression
CO_2 separation unit

Fig. 2.3 Basic VPSA process diagram for CO_2 capturing purpose [21]

Many studies suggest VPSA as a way to separate CO_2 from biogas [22], [23] and from flue gas [21]. In addition, there is one study indicating that the technology is highly effective for flue gas treatment since adsorption equilibrium factors between adsorbate pairs of CO_2/N_2 and CO_2/O_2 are significantly high [24]. The process can recover up to 95% of CO_2 with a purity as high as 99.5% [22]. Normally, the VSPA process runs at 1.35-3.5 bar for the adsorption step and the sub-atmospheric pressure of 0.1-0.05 bar for the regeneration step.

However, there is still limited experience to implement the technology with the post-combustion CO_2 recovery [21] where one of the issues that need further investigation is the influence of presence of water [25].

Monoethanolamine (MEA)

MEA is the most mature process in carbon capturing technology especially for treating natural gas [26]. Absorption reaction occurs at a temperature range of 30-60 °C and operating pressure of 30-50 bar while regeneration of the solvent is at around 107-127 °C and 1.5 bar [18]. This reversible reaction can be described as Eq.(2.5) and the simplified process diagram as Fig. 2.4

$$CO_2 + 2HOCH_2CH_2NH_2 \leftrightarrow HOCH_2CH_2NH_3^+ + HOCH_2CH_2NH(CO_2^-) \qquad (2.5)$$

The capturing is done by introducing flue gas at the bottom of the absorption column while amine solvent is sprayed counter-current to the flow. This allows CO_2 reacts with the solvent and becomes a water-soluble compound [27]. This CO_2-rich compound is then fed to the stripper where the solvent is regenerated and CO_2 is desorbed. This can be done by heating up and decompressing the stripping column. Typically, the concentration of MEA is around 15-25 %wt[18] since higher concentration leads to higher corrosion problems [28]. But by this it is still be able to recover up to 95% of the CO_2 in flue gas while produces 99.5% in CO_2 purity [28].

The process, however, needs very high thermal energy demand for its solvent's regeneration step [21], [28], which is reported to be around 3.3 MJ/kg CO_2 [22]. In addition, the process also has some chemical problems with acidic contaminants like SO_X and NO_X which can form heat-stable or non-reclaimable corrosive salts [26], [28].

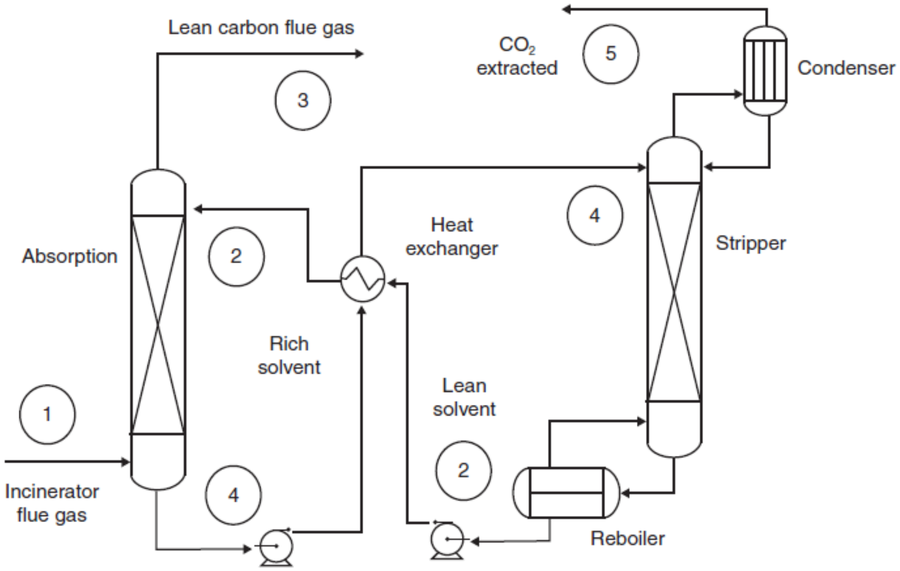


Fig. 2.4 Schematic diagram of MEA-based process [26]

3. Methodology

Technical analysis

This study uses the same size as described in [6] as a base case. The specification of the process is shown in Table 3.1. In order to integrate the process with a waste-to-energy plant, some additional units have to be added to separate CO_2 out of the flue gas.

Table 3.1 Components of the base case power-to-gas process [6]

Component	**Capacity**
Electrolyzer	485 Nm^3/h
Sabatier reactor	1.2 MW of CH_4
District heating connection	1.2 MW
Synthetic natural gas storage	1600 m^3
Oxygen storage	1600 m^3

The waste-to-energy plant

The base case data is adjusted to give some insights for an actual waste-to-energy plant in Sävenäs, Göteborg, Sweden run by Renova. The plant supplies 1346 GWh of heat and 265 GWh of electricity which are approximately 30 and 5 percent respectively of Göteborg's requirements [5]. The key parameters that are used in this study are shown in Table 3.2.

Table 3.2 Key parameter of the waste-to-energy plant [29]–[31]

Maximum energy[1] production capacity	230 MW
Annual operating hours[2] per year	7500 hours
Average flow rate of flue gas	327030 Nm^3/hr
District heating supply temperature[1]	100°C
Chimney temperature	85°C

[1] Including both power and heat

[2] Average value from [30]

Regarding to the energy production capacity, the plant emits 528 ktons (excluding N_2 and O_2) of flue gas in 2013 [29] which the annual amount of several pollutants that are emitted at the chimney between year 2011-2013 can be seen in Table 3.3. Due to strict regulations, levels of hazardous compound like Dioxin, Sulfur oxides (SO_x), Hydrogen chloride (HCl) are significantly low. This is because the flue gas is cleaned by a complex post-combustion gas treatment process which includes electrostatic precipitator, water scrubber, and catalytic converter [30].

Table 3.3 Annual emissions of the plant from 2011-2013 [29]

Compound	**2011**	**2012**	**2013**
Ammonia (ton)	2.8	2.2	1.6
CO_2 (ton)	552800	566800	528000
• Those from fossil-based source (ton)	193400	198300	198400
CO (ton)	69	79	75
NO_x (ton)	179	184	166
Fly ash (ton)	0.6	0.7	0.5
SO_x (ton)	8.6	8.3	9.6
Total organic carbon (ton)	1.8	2.3	2.5
HCl (ton)	0.6	0.6	0.6

Electrolyzer units

As mentioned in Section 2, the AEC is the mature technology for hydrogen production. Hence, it is chosen as a way to supply hydrogen for the Sabatier reaction. The technical data related to the equipment (e.g. proper size, energy demand) are obtained from [6], [32].

CO_2 separation units

Here in this study, two different CO_2 separation alternatives (i.e. VPSA and MEA) are used for the feasibility investigation. They are required to extract sufficient amount of CO_2 in order to feed the Sabatier process to run continuously. According to a data recorded at the waste-to-energy plant, the typical flue gas compositions can be seen as in Table 3.4.

Table 3.4 Waste-to-energy flue gas compositions [29]

CO_2	10.4	%vol dry
O_2	11	
N_2[1]	78.1	
H_2O[2]	20	%vol wet
CO	23.9	ppm
NO_x	34.7	
SO_x	1.46	
NH_3	0.88	

[1] By mass balancing
[2] Average value from [30]

As mentioned in Section 2, it is not optimal to feed the entire flue gas flow directly to the Sabatier process. Thus, a CO_2 separator step is introduced for the waste-to-energy plant. A simplified illustration of the process integration can be seen in Fig. 3.1.

Efficiencies

Efficiencies relevant to the combined Sabatier-CO_2 capturing process are the efficiency of electrolyzer and the process' overall efficiency which both are described as following equations.

The electrolyzer efficiency is defined as,

$$\eta_{electrolyzer} = \frac{Amount\ of\ H_2\ produced}{Energy\ supplied} \tag{3.1}$$

While for the overall efficiency, it is described straight forwardly as,

$$\eta_{all} = \frac{Amount\ of\ SNG\ produced}{Overall\ energy\ supplied} \tag{3.2}$$

Heat potential

Basic chemical process simulation is performed to determine possible heat potential within the process which could be utilized as district heating. Aspen Plus is used for this modelling purpose since the software has database of all the substrate as well as supported ways of calculation.

Economic Analysis

To determine whether the electrofuel technology would be suitable for a waste-to-energy plant, one has to do an economic analysis. Such analysis could show how good the potential and profitability of the process are, in the long run. The cost analysis is based on the economic calculation done in [6]. Some additional costs (e.g. investment on separation units) are, however, added to represent the costs for the specific integration assumed in this study.

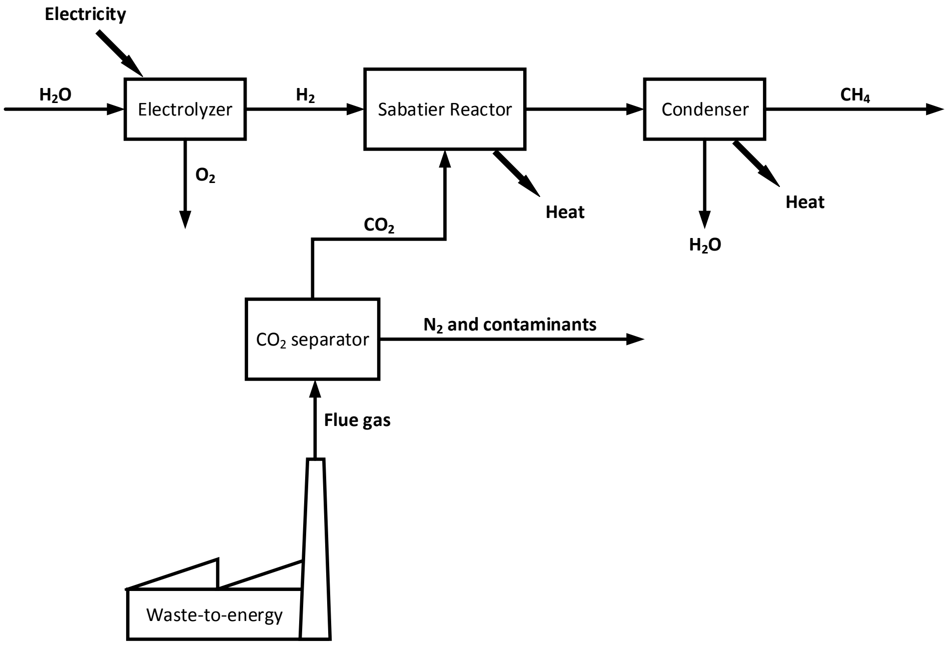

Electricity
H₂O
Electrolyzer
H₂
Sabatier Reactor
Condenser
CH₄
O₂
Heat
CO₂
Heat
H₂O
CO₂ separator
N₂ and contaminants
Flue gas
Waste-to-energy

Fig. 3.1 Power-to-gas process for waste-to-energy plant

Investment costs

When consider size of the case study plant, the costs required only for the Sabatier process are shown as found in [6] while estimated delivered equipment costs of CO_2 separation processes are retrieved from [22]. Due to the fact that the equipment costs are obtained from different period of time, the costs of both processes are then updated to be as of 2014 by using Chemical Engineering Plant Cost Index (CEPCI). The updated costs are concluded as in Table 3.5. When these equipment costs have been summed, the cost for engineering and construction services would then calculate as 52% of the equipment cost [33]. At last, the overall investment cost would be obtained by summarizing the equipment costs with the service costs.

In order to take into account time value of money (TVM) as well as the desired rate of return, the investment cost then has to be converted to an annual cost which is described as Eq.(3.1)

$$Annual\ investment\ cost = TC \times \frac{r(1+r)^n}{(1+r)^n - 1} \tag{3.1}$$

where

TC = Total investment cost

n = Plant lifetime (years)

r = Desired rate of return (%)

For the base case, the plant lifetime is set to be 15 years with the rate of return of 10%. The rate of return is then varied to determine sensitivity of the profitability.

Production costs

The Sabatier process

To produce methane from the Sabatier process without any suspension, utilities like electricity and water has to be supplied continuously. While at the same time, the process also needs the proper produced gas distribution system, catalyst replacement due to poisoning and deactivation, and general operation & maintenance. All of these thus lead to further costs apart from the investment.

For the electricity cost, it is a combination of the electricity price and transmission cost. The electricity price is obtained by averaging the future market price from forward contracts traded in a Nordic market of Nasdaq OMX commodities [34]. The price is set by using averaging prices from the second quarter of 2015 to year 2025. This market price relies on actual prices traded within Nord Pool Spot system as a reference with purpose of price hedging and risk management [35]. The future market price itself is likely to be slightly higher than the actual price [36], but this average price seems reasonable to use as a rough estimate of what can be expected from selling electricity to the grid and therefore could be a reasonable cost estimation for the electricity used for the electrofuel production (where it is assumed that you recycle your own produced electricity). For simplicity reasons we also assume that the waste-to-energy plant can buy additional electricity for this prices as well if more electricity is needed than internally produced.

The average price obtained from the Nasdaq OMX is also checked with another sources [6], [23] to ensure that the price is reasonable. The transmission cost, is estimated from a report published by Swedish Gas Technology Center [23]. To handle the uncertainties, the effect of other electricity prices will be investigated in sensitivity analyses.

Water supply cost is calculated from a median of household water price issued by Swedish Water and Wastewater Association [37]. SNG distribution cost is estimated from [6] where the cost is based on assessment of Swedish Energy Market Inspectorate. For the catalyst price, it is also based on [6] which have been estimated to be 27000 SEK per year.

Table 3.5 Investment cost of the components within the 1.2 MW power-to-gas process [6][22]

Component	**Cost (kSEK)**
Sabatier process	
Electrolyzer	5868
Sabatier reactor	1503
District heating connection	702
Synthetic natural gas storage	1350
Oxygen storage	1350
Total with CEPCI	10663
CO2 separation process with CEPCI[1]	
VPSA	8493
MEA	9976
Engineering and construction services	
Sabatier+VPSA	9961
Sabatier+MEA	10732
Total investment cost	
Sabatier+VPSA	**29116**
Sabatier+MEA	**31370**

[1] The costs are converted to the studied size by using a correlation appeared in [22] as Cost B=Cost A (Capacity B/Capacity A)scaling exponent ; scaling exponent =0.7 for PSA and 0.6 for MEA. The costs were also reported in EUR. Hence in this study, they are converted with an exchange ratio of EUR/SEK =9.

The carbon capturing process

Electricity is needed for running a VPSA process where the details of the requirement can be found in [38]. As the paper reports, the process requires 4-10 kW of electricity per ton CO_2 captured per day expressed in ton (T) per (P) day (D) captured (c) (kW/TPDc). In this study it is assumed that the electricity need is 8 kW/TPDc. Although the cost estimation is based on a post-combustion CO_2 capture for a coal-fired power plant, the flue gas compositions is fairly comparable.

Similarly, for the MEA process, the costs are based on a coal-fired power plant. They can be found in [28]. The paper also mentioned that the process requires roughly 530 kWh of steam, 40 kWh of electricity, and 76 m^3 of cooling water per ton of CO_2. In addition, MEA, caustic soda, and activated carbon have to be added up regularly due to degradations. Hence, expenses of these chemicals are also taken into account.

The total cost of the combined process

According to [6], operation & maintenance cost can be estimated as 5% of the equipment cost (the investment cost without cost of engineering and services). Hence, total cost of the combined Sabatier-carbon capturing process can be calculated as in Table 3.6. The table also reveals the conclusion of all related production costs.

Table 3.6 Production cost of the 1.2 MW base-case combined process

Production cost	**Specific cost**	**Annual cost (kSEK)**
Sabatier		
Electricity[1]	0.278 SEK/kWh	5212.5
Electricity transmission	0.1 SEK/kWh	1875
Water[2]	33 SEK/m^3	87
Methane distribution	0.2 SEK/kWh	1800
Catalyst		27
VPSA		
Electricity	0.278 SEK/kWh	95
Electricity transmission	0.1 SEK/kWh	34
MEA		
Steam[3]	30.8 SEK/ton_{steam}	99
Electricity	0.278 SEK/kWh	20
Electricity transmission	0.1 SEK/kWh	7
Cooling water, 11 °C[3]	0.0315 SEK/m^3	4
Solvent make-up[3]	9 SEK/kg	30
Caustic[3]	3.6 SEK/kg	1
Activated Carbon[3]	27 SEK/kg	2.5
Operation & Maintenance		
Sabatier+VPSA	5% of equipment cost	937
Sabatier+MEA		1007
Total production cost		
Sabatier+VPSA		**10067.5**
Sabatier+MEA		**10172**

[1] As of March, 2015 ; EUR/SEK =9

[2] Estimated from the average median between normal house and apartment prices

[3] The costs were shown in USD; USD/SEK =7

Sale revenue

Main revenue from the electrofuel process is SNG. Hence, its price is a key parameter in the cost analysis to determine if the process is profitable or not. The price is selected based on data reported in [6], [39]–[41]. A sensitivity analysis is also performed to check how sensitive the annual profit is due to the SNG price.

Important byproducts from the electrofuel process are district heating and oxygen. Since the process emits excessive heat and the waste-to-energy plant itself has an existing district heating supply system, it is advantageous for the plant to utilize such heat for the public district heating network. The revenue earned from the heat is estimated from a sustainability report [42] published by the case study waste-to-energy plant using annual income of the plant together with the estimated electricity price. The resulted price is then 0.171 SEK/kWh. For the oxygen, since the electrolysis yields very high purity oxygen, it can be used for any chemical purposes. Its price is set as 0.51 SEK/kg found in [43]. Initial values for the plant's income sources are listed in Table 3.7.

Sensitivity analysis

The combined process is varied in size from the initial base case size of 1.2 MW. By using data from [29], two maximum possible sizes are determined: Case A, the size adjusted to the amount of electricity for sale (produced in the waste-to-energy plant) and Case B, the size adjusted to the amount of CO_2 produced in the waste-to-energy plant, more specifically based on the lowest flow of CO_2 emission that could serve the process to run continuously without any CO_2 supply shortage. The reason for the first choice of size is to investigate if the process is profitable when total electricity, which otherwise sold to the grid, is utilized solely for the gas production purpose.

To upscale the size and to take into account the economics of scale, Eq.(3.2) is introduced for the investment cost calculation [44]. The sixth-tenth rule (k=0.6) is assumed for the Sabatier and the MEA process while for the VPSA, k=0.7 is used for the change in scale of the unit. More detail about the k-exponent for CO_2 capturing process can be found in [22].

$$\frac{C_a}{C_b} = \left(\frac{A_a}{A_b}\right)^k \tag{3.2}$$

where

C = Investment cost

A = Capacity of the process

k = Cost exponent

and subscripts

a = Desired capacity

b = Known or base capacity

The proportionally-increased production costs and revenues are calculated as in the initial case before every economic aspect will be summed up to determine the total annual profit of each case.

The three cases, the base case, Case A and Case B, are then analyzed for sensitivities against the annual profit of the combined process. Main process parameters that are varied are: SNG price, electricity price and change in investment costs. The results could show ranges of prices and costs that are break-even for the different process sizes

In Case B, the electricity required for the gas production is much higher than the electricity the plant can produce. The plant therefore needs to purchase electricity from external sources. In such scenario, cost of electricity could be higher than that of the self-produced since external electricity provider could add some margin from selling the electricity while, on the other hand, using electricity for gas production could also be entitled to pay relevant taxes. Hence, to take into account these uncertainties, changes in cost of electricity in form of percentage from the initial price and projected prices are applied to determine the outcome.

Another factor that could possibly be critical is electrolyzer's efficiency. Hence in this study, the efficiency is varied between ranges of 40-90% to determine the influence on the gross profit.

Table 3.7 Possible revenues earned from the base case 1.2 MW power-to-gas process

Source of income	Specific income	Annual income (kSEK)
Methane	1.48 SEK/kWh	13320
Oxygen	0.51 SEK/kg	1188
District heat	0.171 SEK/kWh	1724

4. Results

Sizes of Case A and Case B

According to the plant's report, the annual amount of electricity sold to grids is 189 GWh per year. By dividing the amount with annual operating hours, it corresponds to the electrofuel process size in Case A as 11.9 MW, or nominally 12 MW, SNG production capacity. For Case B, based on the reported flue gas emission, mass-basis CO_2 emission can be obtained as shown in Fig. 4.1. The emission is lowest in June, which is the month that is used to determine the electrofuel process size in Case B. Adjusting the process to June's emission results in a 241.5 MW (or nominally, 240 MW) SNG production capacity.

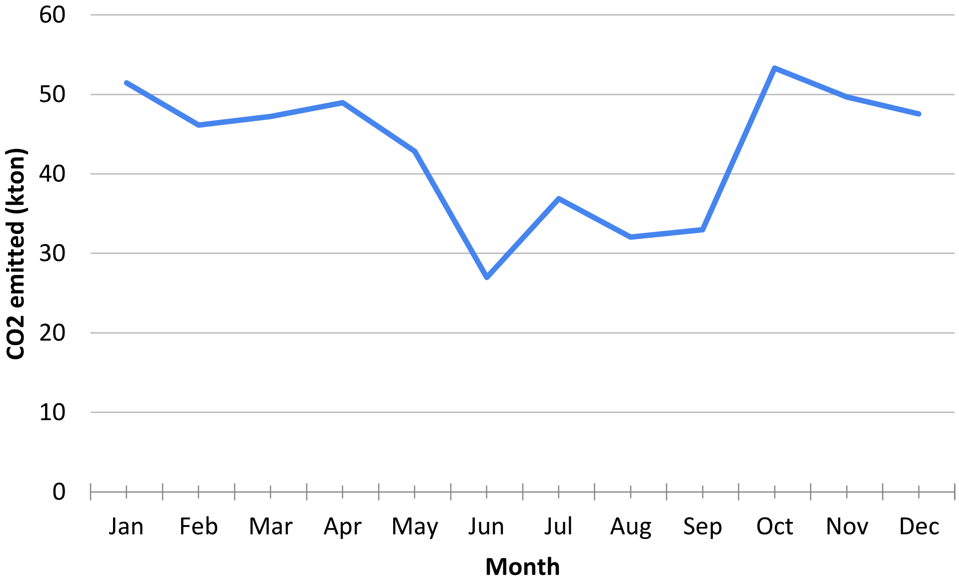


Fig. 4.1 Monthly CO2 emission from the wate-to-energy plant in year 2013

The preliminary results

By using the primary production costs and revenues as stated on the Methodology section, the total investment cost, operating cost as well as profit of the three different cases can be seen as in Table 4.1.

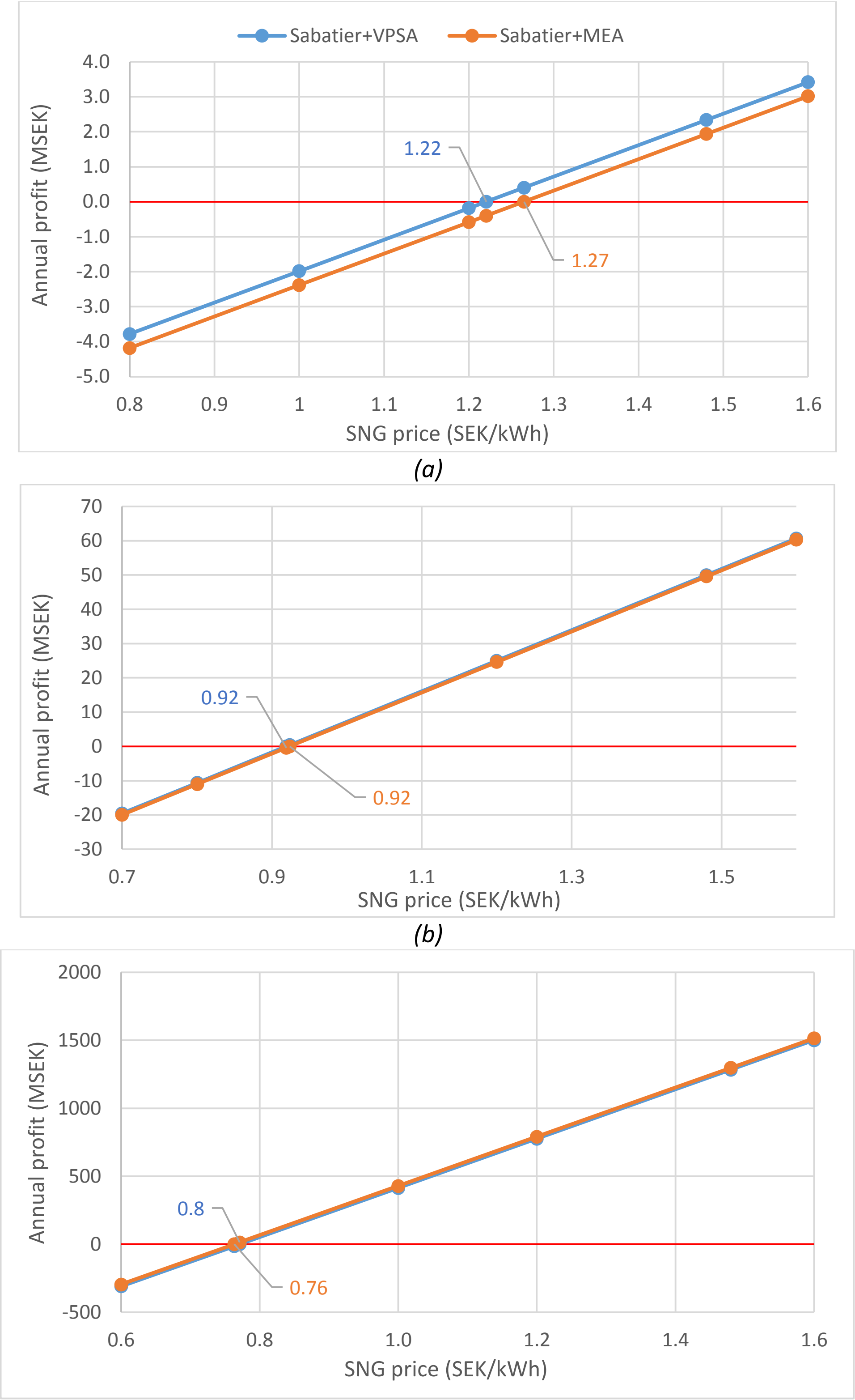
The results show that with the process size of 1.2 MW, the annual profit of the combined process of Sabatier and VPSA is slightly higher than that of Sabatier with MEA. This is due to both higher investment and operating costs of the latter one. The same trend happens with Case A, although the difference is even smaller. However, by an increasing influence of the economics of scale, the larger plant size of Case B shows that the Sabatier with MEA has a higher annual gain even though the operating costs of the process is still higher than its counterpart. The efficiency of overall production is 47 % for Sabatier+VPSA and 48 % for Sabatier+MEA in every case. While considering only the electrolyzer, the figure is 69 %.

Table 4.1 Preliminary results in million SEK per year.

	The base case (1.2 MW)	**Case A (12 MW)**	**Case B (240 MW)**
Annualized investment cost			
Sabatier+VPSA	3.8	16.3	116.0
Sabatier+MEA	4.1	16.3	99.5
Annual operating cost			
Sabatier+VPSA	10.1	94.5	1866.7
Sabatier+MEA	10.2	94.8	1869.3
Revenue earned			
CH4	13.3	131.9	2681.0
O2	1.2	11.8	239.2
District heat	1.7	17.1	346.9
Annual profit			
Sabatier+VPSA	2.3	50.0	1284.4
Sabatier+MEA	1.9	49.6	1298.3
Annual profit in rounded values			
Sabatier+VPSA	2.3	50	1280
Sabatier+MEA	1.9	50	1300

Effects of SNG price

It is clear as in Fig. 4.2 that the variation on SNG price plays the most significant role for the combined processes to get profitable. The higher price you can sell the produced SNG for, the better, and consequently the lower price you need until you reach a break-even point, the better. Break-even point of the SNG price decreases as the plant size expands. For example, the break-even SNG price for the base case Sabatier+VPSA is 1.22 SEK/kWh while for the Case B, the figure is declined to just 0.8 SEK/kWh. This is due to the effect of the economics of scale on the investment cost that makes the profitability higher, with a larger electrofuel process, since the overall capital cost per unit of SNG is lower in the larger plant capacity. The Sabatier with MEA has a slightly higher break-even SNG price in the base case and Case A since the investment cost of the process is still higher than the Sabatier with VPSA. In Case B, however, the tolerable price of SNG is lower for the Sabatier with MEA by a difference of 0.04SEK/kWh SNG from the Sabatier with VPSA. The difference in the break-even price between configurations in Case B is actually higher than that of Case A since deviations in investment cost between the two combined processes is greater.



(a)

(b)

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Fig. 4.2 Effect of SNG price variation on annual profitability of (a) the base case, (b) Case A, and (c) Case B. The numbers in blue and orange indicate the break-even price of Sabatier+VPSA and Sabatier+MEA respectively. Note that the axis scale differs between the graphs.

Influence of electricity price

Similarly to the SNG price variation, the tolerable electricity prices are affected by the economics of scale of the larger capacity, which is due to the fact that the proportion of the investment cost within the total cost is considerably reduced with the larger plant size. Hence, change in any operating expense like cost of electricity would be greatly influential.

The Sabatier with VPSA combined process is less tolerant to the change in electricity price since the process consumes more electricity than the Sabatier+MEA. The lower tolerable electricity price is expected in Case A and B, but for the base case the trend is reverse because the investment still has a large share in the total cost. Trends of changes in annual profit by electricity price variation can been seen as in Fig. 4.3

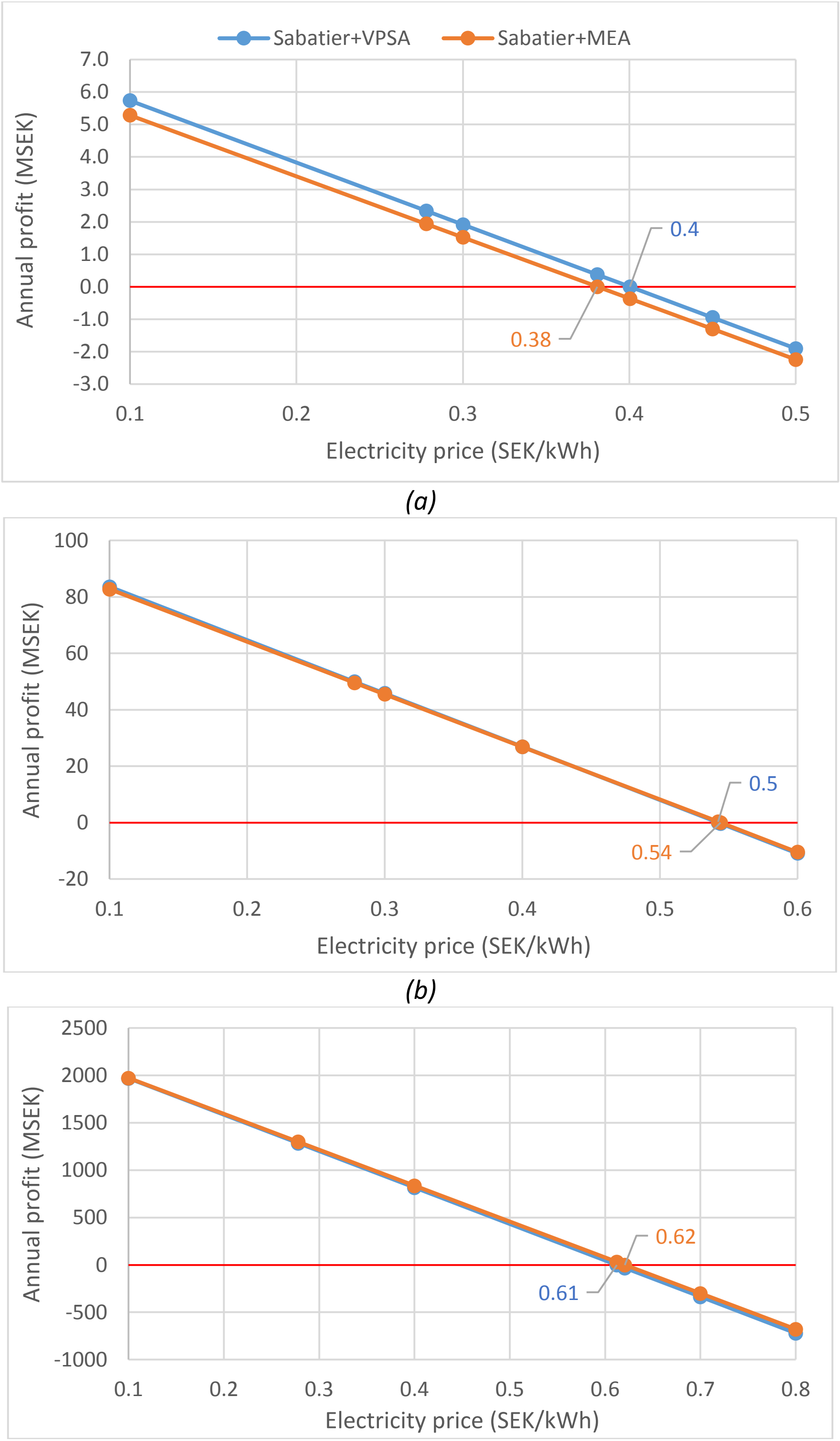
Since the plant can produce only 189 GWh of electricity per year and the power requirement for Case B is roughly 3.9 TWh per year, it means that generated electricity is insufficient and external electricity is thus needed.

When the price of internally-produced electricity is set to the initial value of 0.278 SEK/kWh, results as in Fig. 4.4 show that both in Case B are profitable until the change in electricity price reach the break-even line at 126% and 130% for Sabatier+VPSA and Sabatier+MEA respectively.

Change in investment cost

The total investment cost

Since there are some uncertainties about future investment costs, especially regarding the electrolyzer, the effect from changing the total investment is also investigated. When the total cost is varied between −40% and 120% (negative percentage represents a less costly investment) as in Fig. 4.5, the only loss gained from all the cases is in the base case when the cost is more expensive than the initial value by 61% or 47% for Sabatier+VPSA or Sabatier+MEA respectively. Less effects on change in the cost could be seen in the processes with higher production capacities. An example for this is Case B, when the investment cost is higher than the starting point by 120%, the annual profit reduces by only roughly 10%. This is again due to the economics of scale which makes the process cost having a lower share of the total investment cost per unit of SNG produced.



(a)

(b)

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Fig. 4.3 Effect of electricity price variation on annual profitability of (a) the base case, (b) Case A, and (c) Case B. The numbers in blue and orange indicate the break-even price of Sabatier+VPSA and Sabatier+MEA respectively. Please note that the y-axes are not identical in different cases

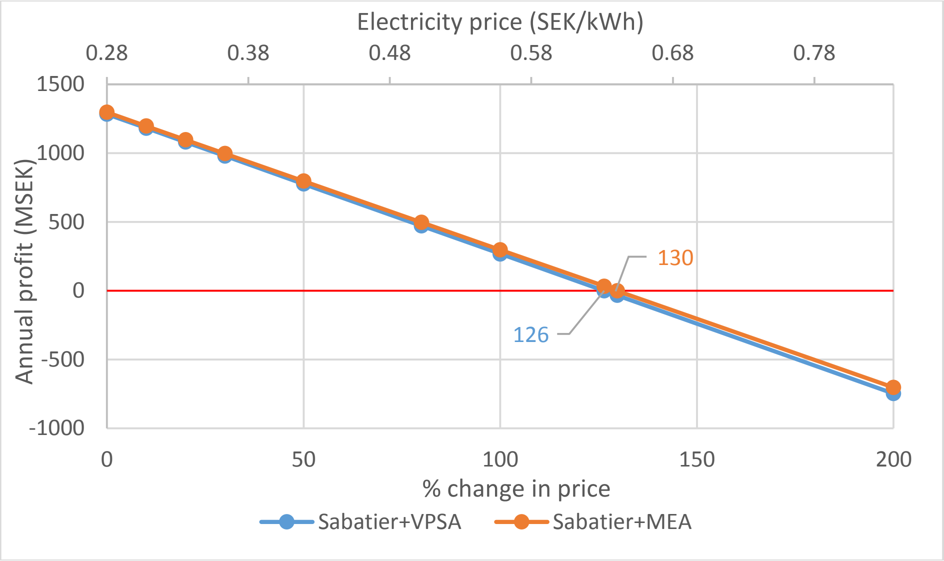


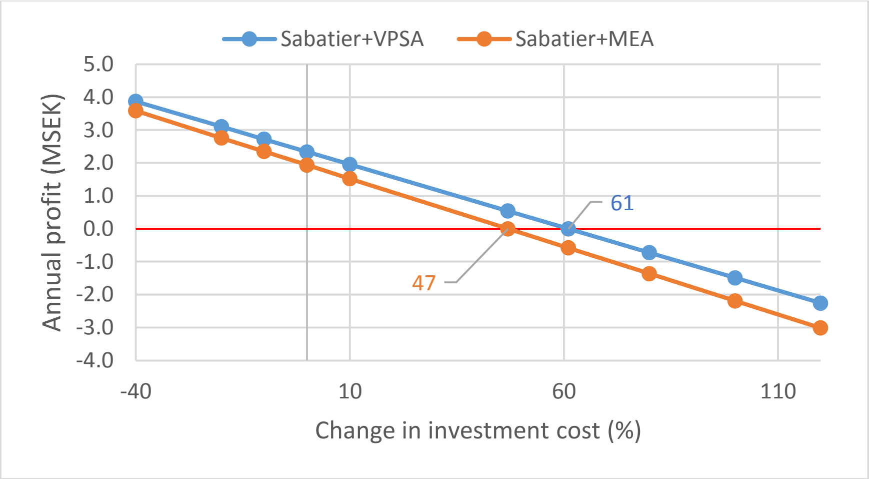
Fig. 4.4 Effect of externally-purchased electricity price on annual profit of Case B. When self-produced electricity price is initially set to 0.278 SEK/kWh.

Since the Sabatier process consists more or less of the basic chemical process equipment, its cost can be estimated straight forwardly with the cost evaluation which has been mentioned in the Methodology section. However, for the processes like electrolyzer or VPSA which contains some special equipment that can be difficult to estimate in a simplified way. Therefore, additional investigations have been taken place to consider the variation on the cost specifically for the electrolyzer and CO_2 capturing process. From the literature it is reported that the cost of the electrolyzer unit can be as high as 1.5 times more than assumed in the initial case [23]. For the CO_2 capturing process, it is still unclear on what should be a proper range of the investment cost since the size of process used in the study is small. It is possible that the cost could be much higher than assumed as initial value. Thus, range of cost variation is selected for both electrolyzer unit and CO_2 capturing to be between 0% and 400% to determine their effect on the annual profit.

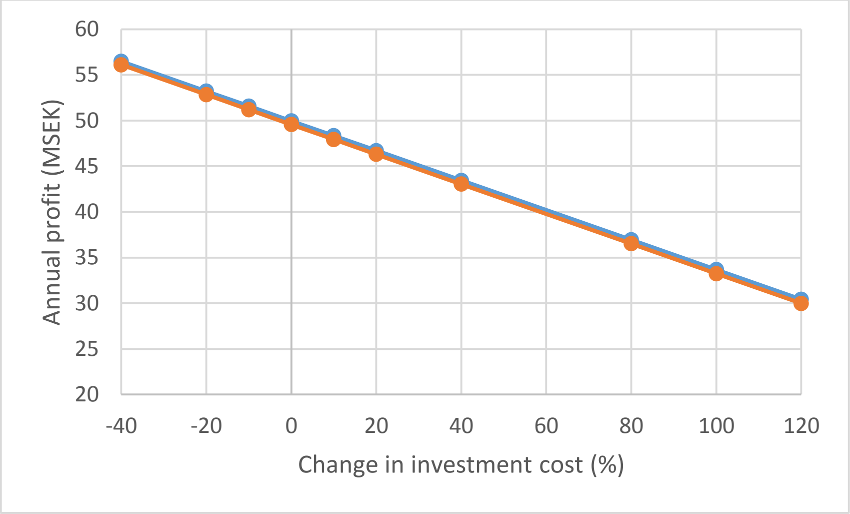
Results show that in the base case of 1.2 MW, the cost of electrolyzer cannot exceed 201% from the original price in case of Sabatier+VPSA and 167% for Sabatier+MEA until the production become unprofitable. However, in Case A and Case B, the plant is still profitable even though the price on electrolyzers are increased by 400%.

For the CO_2 capturing, the same trend can be obtained. For the base case, the break-even percentages are slightly lower than that of the electrolyzer: 97% and 138% for Sabatier+VPSA and Sabatier+MEA respectively. The trends of both variations can be seen as in Fig. 4.6.

Besides, there is also a scenario when both processes are together more expensive than the initial evaluated cost. To study such case, the investment cost of the electrolyzer is assumed to be 150% higher than the initial cost. The CO_2 capturing is then varied within a range of 0-400%.



(a)



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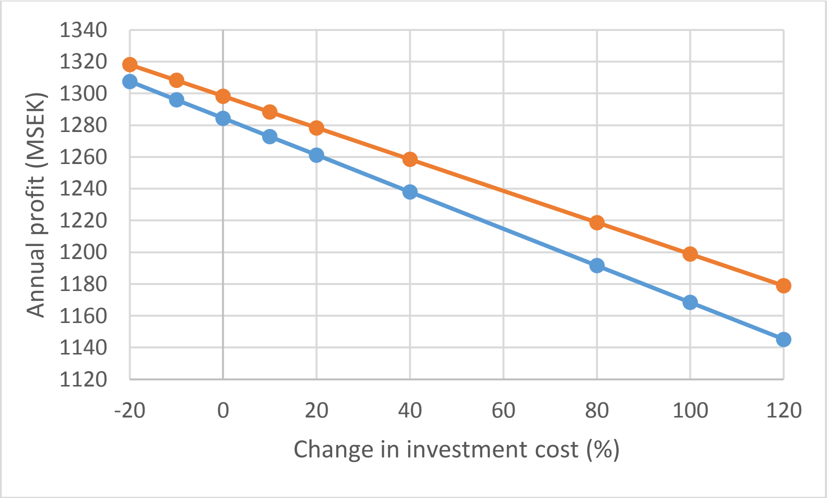
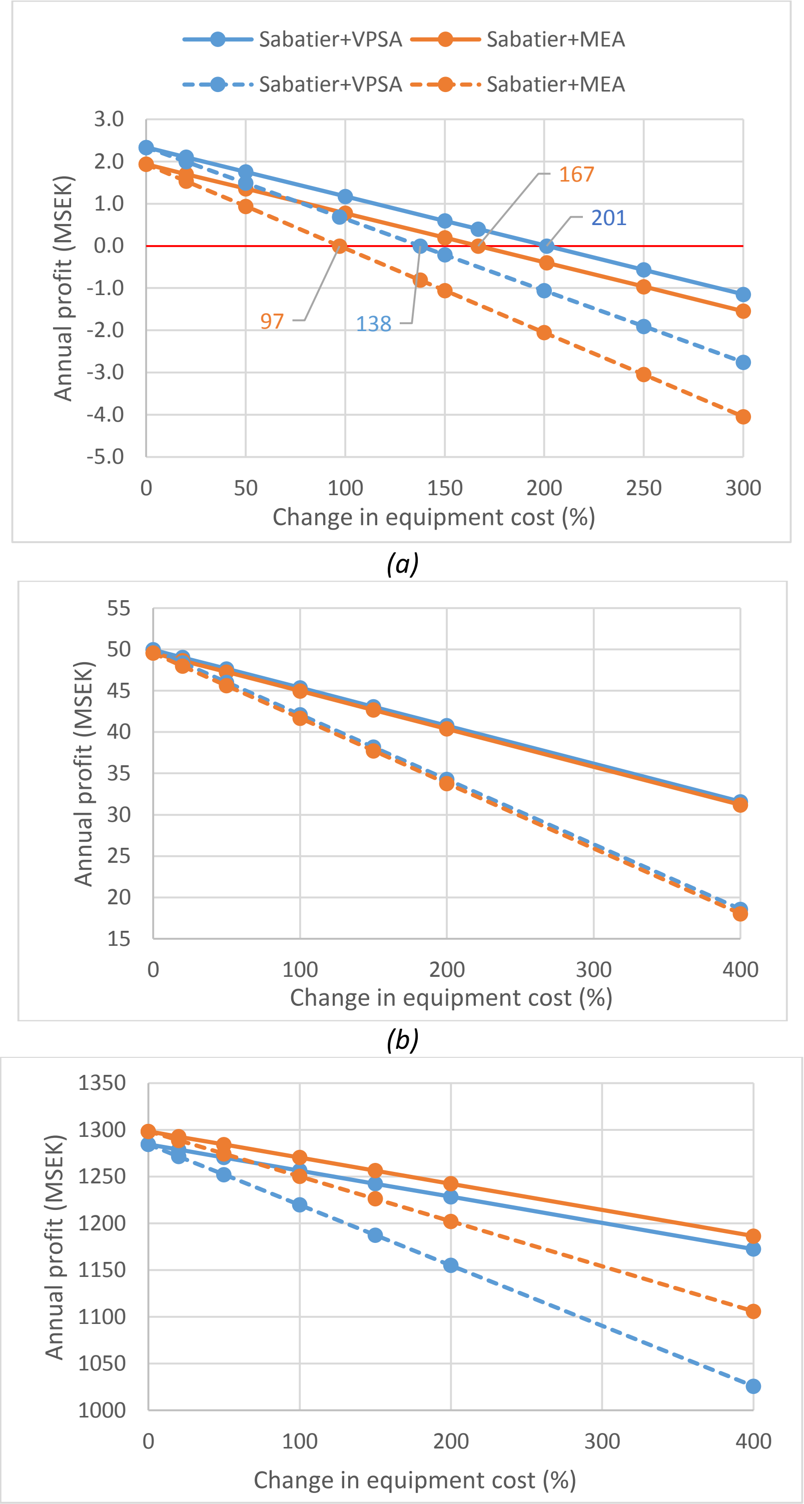


(c)

Fig. 4.5 Effect of total investment cost on annual profitability of (a) the base case, (b) Case A, and (c) Case B. The numbers in blue and orange indicate the break-even price of Sabatier+VPSA and Sabatier+MEA respectively. Please note that the y-axes are not identical in different cases.



(a)

(b)

(c)

Fig. 4.6 Effect of investment cost of (—) electrolyzer and (- - -) CO_2 capturing process on annual profitability of (a) the base case, (b) Case A, and (c) Case B. The numbers in blue and orange indicate the break-even price of Sabatier+VPSA and Sabatier+MEA respectively. Please note that the y-axes are not identical in different cases.

The results as in Fig. 4.7 illustrate that the profitability is vulnerable when the cost of CO_2 capturing process is higher than the initial set point by 11.5% for the base case 1.2 MW Sabatier+VPSA plant. For the Sabtier+MEA, the process has already been unprofitable before the cost is changed. This is because the higher investment cost of the electrolyzer and its H_2 storage has let the process to gain loss. The SNG production remains profitable for the larger capacities even though the annual return is significantly reduced.

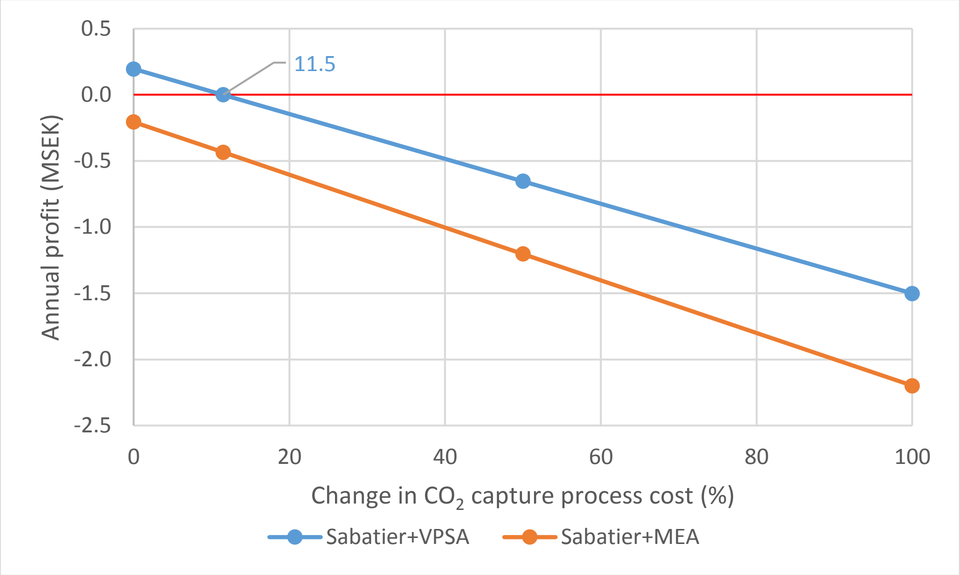


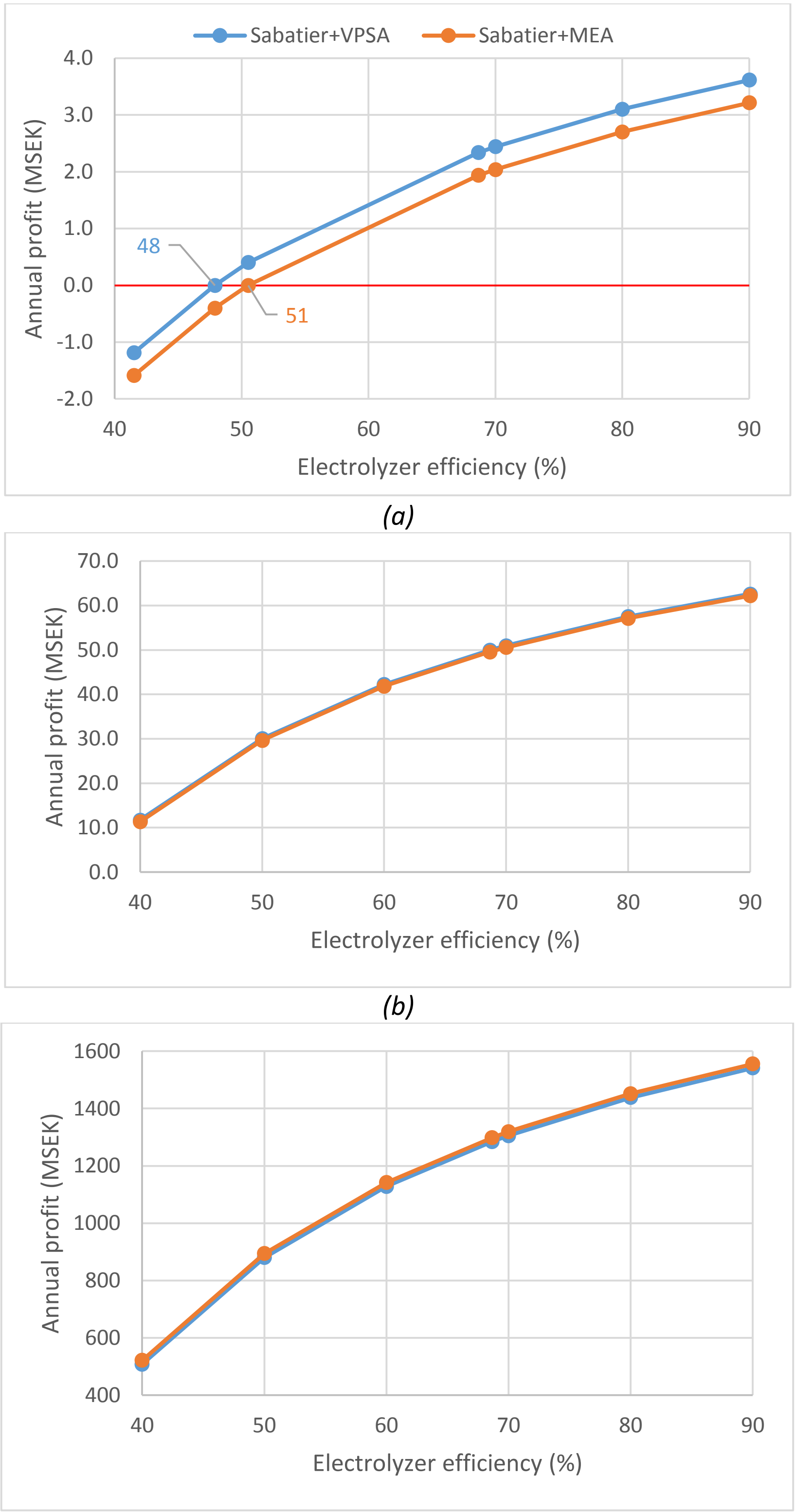
Fig. 4.7 Effect of investment cost of CO_2 capturing process on annual profitability of the base case, assuming a fixed higher investment cost of the electrolyzer to 150% higher than the initial cost. The numbers in blue indicates the break-even price of Sabatier+VPSA.

Effects on variation in electrolyzer efficiency

As shown in Fig. 4.8, the efficiency of the electrolyzer plays significantly role to the profitability of the whole system. Electrolyzers with a lower efficiency means more electricity has to be supplied to produce the same amount of H_2 and results in a higher operation cost. For the base case, the efficiency can reach down to 48% in case of Sabatier+VPSA and 51% for the Sabatier+MEA before the electrofuel production will get unprofitable. For the larger capacities, the process remain profitable even when the efficiency of the electrolyzer is decreased to as low as 40%, however, considerable decline in profit can be observed. For example, in case of 12 MW Sabatier+VPSA, the profitability of the electrolyzer with 40% is 5 times lower than that of the initial value of 69% and almost 6.5 times lower in case of 90% efficiency.

Plant size change due to seasonal energy production variation

As the waste-to-energy plant is intended to generate district heating as its main product, the plant can earn much of its revenue during winter time when there is a high demand of heating. In contrast, the plant earns much less income during summer due to a very low demand. Therefore, the plant normally has a maintenance period during the summer months. It also means that the plant’s energy production would decrease and could affect continuity of electricity supplied to the combined process.



(a)

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Fig. 4.8 Effect of electrolyzer efficiency on annual profitability of (a) the base case, (b) Case A, and (c) Case B. The numbers in blue and orange indicate the break-even price of Sabatier+VPSA and Sabatier+MEA respectively. Please note that the y-axes are not identical in different cases.

According to the plant information [31], the electricity production as illustrated in Fig. 4.9 is significantly dropped during summertime as a result of lower energy production capacity. Another issue is that the district heat price could be declined sharply due to the low requirement. The plant determines the period between May and September as summertime, while the rest as winter.

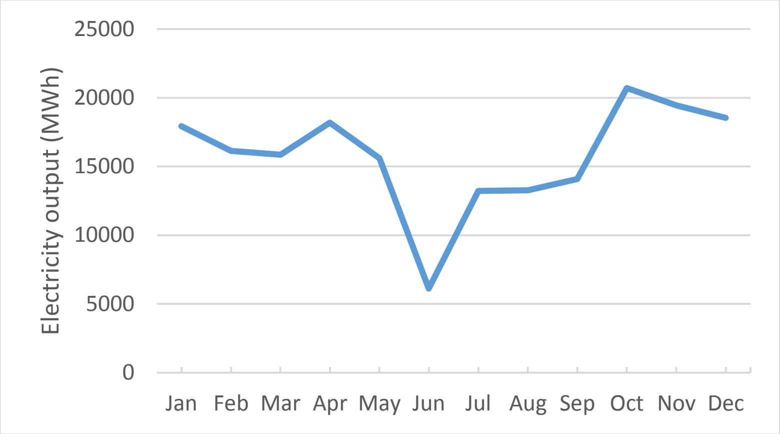

Fig. 4.9 Monthly electricity output of the waste-to-energy plant

The combined process size of Case A, as it is designed based on the yearly-average electricity output, would definitely be affected by such electricity shortage. Hence, to take into the account the influence of seasons, one needs to redesign the Case A into two different scenarios: the summer case (Case A.1) and the winter case (Case A.2).

In summer case, the lowest monthly electricity production between May and September that allows the combined process to receive constant electricity is used to run a continuous methane gas production. While in the winter case, the criteria is to choose the lowest monthly electricity production during winter to let the process to run without pause during winter and with proportionally lower capacities during summer. The criterion are illustrated as in Fig. 4.10

To determine the adjusted electrofuel process sizes, the monthly amount of electricity produced is divided by the operating hours of the month. To avoid oversizing, the maximum operating hours is used instead of the average value to obtain the smallest sizes of the plant. Then, the result is converted to the SNG production capacity by considering the overall efficiency from the preliminary results section. After that, the values yield the monthly amounts of both SNG and district heat produced by multiplying the production capacity with average operating hours. Calculations using the average hours would illustrate that any events (e.g. maintenance) within a month is automatically included. The district heat price during summer is assumed to be 90% lower than that of the rest of the year.

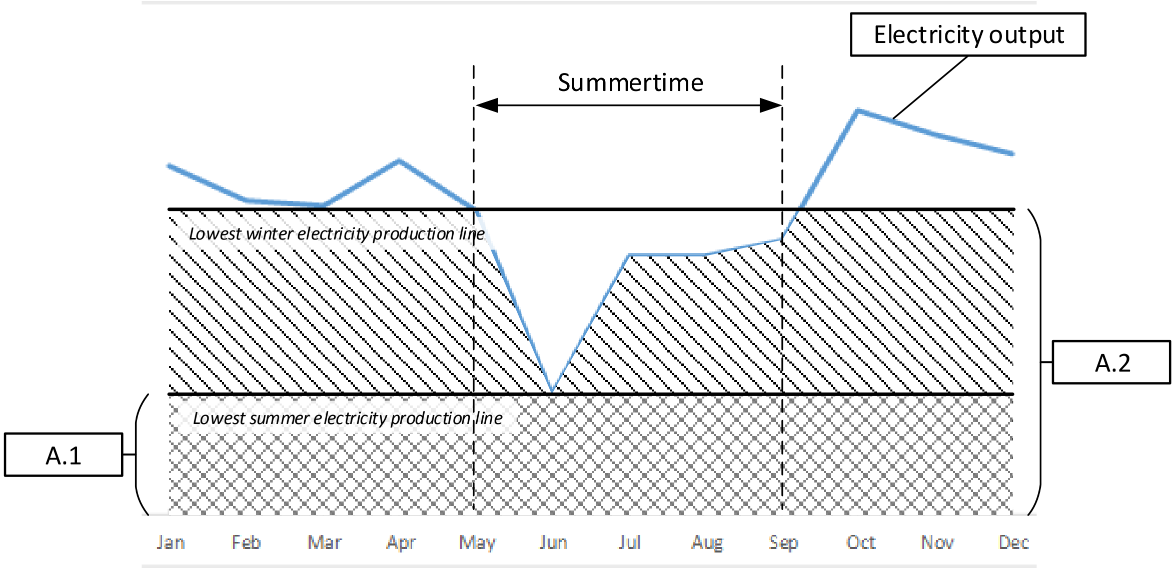

Fig. 4.10 Criterion for the two different A-cases. Case A.1 uses the lowest summer electricity production line as a basis for calculating the plant size whereas Case A.2 are based on the winter line.

By using such criterion, results can be obtained as shown in Fig. 4.11 and Table 4.2. The winter case has an electrofuel process size of 10.1 MW based on the electricity production in March. In summer, the plant is forced to lower down its SNG production to as low as the electricity production in June, leading to an electrofuel process size of 4.6 MW. Note that neither the winter case nor the summer case comes up to an SNG production capacity of 12 MW which was the size assuming annual average electricity production. It can be observed that even though the size of Case A.1 is just more than double compared to that of Case A.2, its total profit is almost 5 times higher. This is due to the fact that the winter case can generate more SNG and much more byproducts throughout a year.

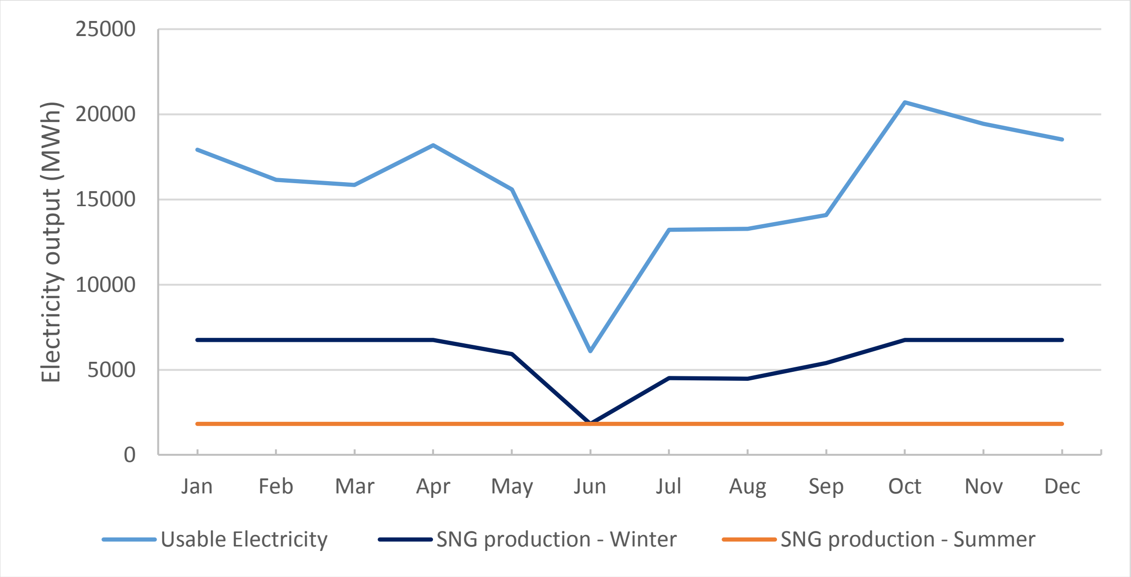

Fig. 4.11 Monthly usable electricity and corresponding SNG produced of two different A-cases

Table 4.2 Economic results of Case A.1 and A.2

	Case A.1 (Winter)	**Case A.2 (Summer)**	
Plant size	10.1	4.6	MW
Annualized investment cost			
Sabatier+VPSA	13.8	6.6	MSEK
Sabatier+MEA	14.0	7.0	MSEK
Total operating cost			
Sabatier+VPSA	73.9	23.8	MSEK
Sabatier+MEA	74.2	24.0	MSEK
Annual revenue earned			
CH4	102.7	32.3	MSEK
O2	9.2	2.9	MSEK
District heat	9.5	2.6	MSEK
Annual profit			
Sabatier+VPSA	33.6	7.4	MSEK
Sabatier+MEA	33.1	6.8	MSEK

Revenue gained from district heating in these different cases is represented in Fig. 4.12. Both cases generates very low income during summer when June is the only month that the revenue is the same in both cases. While in the greater period of a year, the winter case earns almost 4 times higher compared to the summer one.

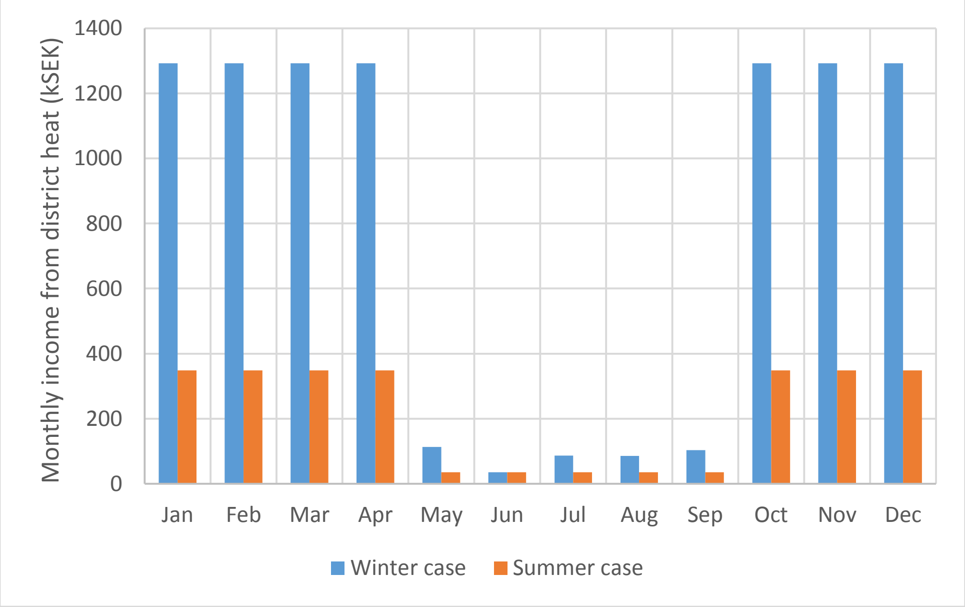

Fig. 4.12 Revenue earned from district heat in two different A-cases

5. Discussion

SNG price

There is no doubt that the SNG price is a determining factor that judge whether the CH_4 production is profitable or not. The initial price used in this study can be very optimistic since it is based directly on the Swedish vehicle gas price which all the relevant taxes are included. The SNG price could, on the other hand, be similar to that of natural gas (NG) which is logically cheaper than the electricity (since NG often are used as feedstock for electricity production). If such price is applied, the process will be unprofitable. Hence, there are some complications on what SNG price can be assumed for this electrofuel production.

However, as the proportion of renewable source supplied to the waste-to-energy plant is as high as 60%, it is reasonable that the price lies closer to that of bio-methane rather than that of fossil natural gas or that 60% can be sold as bio-methane and 40% as fossil methane. By assuming the latter, the plant's revenue might still be able to cope with all the expenses and gain some profit from the production. This, however, is applicable only for the large production capacity where the break-even SNG price is low. For example, according to Eurostat [45], the natural gas price including taxes and levies in Sweden was approximately around 0.476 SEK/kWh in 2013. If the vehicle gas price of 1.48 SEK/kWh is applied, the criteria would give the SNG price of 1.08 SEK/kWh. It means that Case A and B could still be able to cope with all the expenses and be profitable in this scenario.

It is also very likely that the SNG gas price by itself will increase within the period of plant's life time since the price is closely related to prices of natural gas and petroleum (which since they are limited resources are affected by scarcity rent and thus increase in price when the demand is higher than the supply). Data from Eurostat [46], [47] indicates that the natural gas price historically has rapidly increased which, in some period, has been more than 3 times higher within just seven years. Hence, there is a very strong possibility that the SNG price will follow the trend and may be sold to higher prices in future. The concept could be very promising also due to high governmental supports towards greener energy.

Electricity price

The price represented in the calculation is an average price from future trading. Although the same price range has been used in many sources, the price can be varied dramatically in the waste-to-energy case. The price can be much cheaper than the evaluated number since it may receive some additional waste-handling revenue from both local customers and contracted countries. In such case, integration of flue gas methanation system would be highly favorable. The plant, in contrast, may also operate under very unprofitable circumstance due to its complex waste management and treatment which means it needs great supports from either government or owning municipality.

There is some possibilities that the electricity used for gas production within the waste-to-energy plant could be entitled to pay a tax which could be as high as 0.32 SEK/kWh [31]. However, when considering the origin of electricity used in different cases, self-produced electricity should be relieved from any tax. This is due to the fact that the electricity that are normally sold to the grid is now changed its purpose to be used internally instead. The plant may nevertheless need to modify its product description by adding SNG as one of its output products, to be exempted from any possible tax.

In the other hand, in some scenarios like in Case B which some external electricity has to be supplied to run the gas production, implementing such combined process could be entitled to pay electricity tax. If the tax is 0.32 SEK/kWh, then the electricity price is 115% more expensive compared to the base price. Although the process is still in a profitable interval according the sensitivity analysis, it is hard to say that the process will be brought to use since it could have less attractiveness.

Efficiency of the electrolyzer

As mentioned in previous chapter, the efficiency of the electrolyzer affects significantly the profitability of the process. It also implies that invest in efficient electrolyzer would make the whole process more economically favourable in the long run. This is due to the fact that operating cost has a very high proportion in the annual total cost especially for the larger plant size. Hence, less electricity used in water electrolysis means not only lowering in operating cost, but also to make the process potentially less vulnerable to some key price variations (i.e. SNG price and electricity price).

Although this study uses AEC for the electrolysis, one possible way to increase the efficiency of the process is to change the electrolyzer technology to SOEC. The technology uses steam instead of liquid water as its feedstock. Therefore, it needs much higher operating temperature of around 500 to more than 800 °C [2], [48]. However, the efficiency of such alternative is significantly higher than the AEC up to around 90% [2] which the sensitivity analysis on electrolyzer efficiency indicated had a strong effect on the profitability of the combined process. In addition, electricity requirement to produce H_2 is also much lower compared to the AEC. The electricity demand is around 4.3-5.5 kWh/m^3 H_2 for the AEC while for the SOEC, it needs less than 3.2 kWh to produce the same amount of H_2. Hence, the technology seems promising and could be very beneficial to the combined process. Currently, the technology is being tested at Sunfire facility in Dresden, Germany[48] where it is claimed to have the electrolysis efficiency of higher than 90%. This thus indicates that the technology is now marketable and could be extensively used in near future.

Investment cost

For the investment cost investigated in this study, one has to be aware that the cost is taken from only two sources (i.e. Mohseni et al [6] and S. Heyne and S. Harvey [22]) which means that there could be some uncertainties around the cost. Hence, there might be more precise cost-analyses published that can either verify this study's assumptions or propose a new way of the cost calculation.

As in the Results section, the share of investment cost in total annual cost decreases as plant size expands. However, this is due to the economics of scale which costs of all units are roughly estimated using simple exponent. When such an exponent is higher, the investment cost would get a higher share in the total cost. At the maximum point when the exponent is equal to 1, the economics of scale will has no influence to the investment cost of either the whole process or a specific part. Thus, the investment cost of the larger plant size would proportionally be the same with that of the base case of 1.2 MW. This causes the break-even prices of both SNG and electricity as well as tolerable value of other variations will be exactly the same as of the base case.

At least one possibility that the economics of scale could not be applied is suggested in [23]. It reports that the methanation reactor is likely to have the exponent equal to 1. However, if the process has been developing in near future, the investment cost of every single units in the process can get considerably cheaper.

Another possible issue related to the investment cost is due to the fact that in cases like Case A, A.1 and B are able to experience some difficulties trying to lower down their rate of production during summertime. Since the investment cost estimation of the larger plant is based on such a small plant size of 1.2 MW, it could imply that the up scaling faces unforeseen difficulties. For example if the larger process consists of only one single gas production unit it might be difficult to change between winter and summer capacity. In worst case, the plant has to be completely shut down during summer which could cause considerable loss in revenues. If difficulties scaling up the electrofuel production process capacity, one solution may be to install many small production units in parallel. By that, the capacity can be adjustable throughout a year but also means that much more capital has to be invested in.

Plant size

As being described before, the considered waste-to-heat plant can produce roughly 189 GWh/yr on average of electricity, it could not supply the very high demand of electricity in case of the largest electrofuel plant capacity of 240 MW where as high as 3.8 TWh/yr of electricity is required. External electricity has to be introduced in order to handle the need. Additional renewable electricity would add to the positive image of producing renewable methane, whereas it would be less beneficial if the origin of such electricity comes from non-renewable sources. Most beneficial, from a cost perspective, would of course be if the additional electricity could come from excess electricity that otherwise would have been wasted. Such situation could be more common in future if more solar and wind power are introduced in the electricity mix. When these intermittent (weather dependent) sources produce more electricity than demanded this excess electricity is typically almost free of charge. Since the CO_2 emitted from the waste-to-energy plant is excessive, the plant could be designed in any size depending on how large the renewable-based excess electricity is. Future work could focus on finding such size in order to make the most use of generated electricity as well as to assess this electrofuel implementation as an alternative for electricity storage.

Heat recovery

Since, flue gas temperature at the waste-to-energy plant's chimney is around 85°C. The flue gas would then be able to preheat the upcoming feeding-to-electrolyzer water. However, since the water requires to be at the operating temperature of around 80°C, some external heating still need to avoid excessive heat exchanger size and related investment cost due to very low heat transfer driving force between streams.

From a district heating point of view, another possible heat source for supplying heat is at the effluence of the CO_2-free flue gas at the MEA CO_2 capturing process. The temperature of such stream is around 107-127°C [18] which means that a relatively high amount of heat can be utilized for district heating.

An alternative output

Besides applying Sabatier reaction to produce methane from the plant's flue gas, another possibility is to instead using methanol synthesis which basically utilizes the same reactants. The reasons are because the reaction requires less hydrogen flow compared to the Sabatier (Eq.(4.1)) and its product as methanol is easier to manage since it can conveniently be stored in liquid form.

$$CO_2 + 3H_2 \leftrightarrow CH_3OH + H_2O \qquad \Delta H_R^0 = -49.43\ kJ/mol \quad (4.1)$$

The reaction, however, has a very low yield per pass of around 10% [49] due to generation of water that highly inhibits the activity on the surface of catalyst. It may, however, be possible to recycle the components that did not react in the first pass several times until the yield has increased. There is in any case a trade-off between lower in electricity cost and decrease in production yield.

The VPSA performance

Although being a technology of choice for capture CO_2 in the biogas process which normally contains 35 %vol of CO_2, there is still unclear whether the technology could be capable to extract CO_2 from such a low concentration like in this case. One study [38] indicates that capture cost in its considered conditions can be much higher when concentration of CO_2 is low. It also points out that the capturing performance of the process could be worse with low concentration of CO_2. It means that the process may need to be larger in order to handle such low concentration. Hence, some penalty on the process's efficiency could be inevitably affected [50]. Further studies are thus needed to find both optimal pressure range for the low-CO_2-content flue gas as well as better adsorbents.

Sustainability perspectives

SNG produced by implementing the electrofuel technology could be used as an alternative for more renewable transport fuel. It also implies itself as one of the promising ways to replace fossil fuel dependency which in case of waste-to-energy plant, such replacement done by improving energy extraction from waste. This as well could lead the country to improve its energy security and become less dependent on imported fuels.

For a clearer view, if all the waste-to-energy plants in all of 290 municipalities in Sweden decide to apply the technology with just the summer case's size of 4.6 MW, the amount of gas produced could substitute 860 kiloton of oil which contributes about 11% of 2011 total energy requirement in transport sector including shipping and aviation (or 13% of road transport)[51].

From the plant point of view, it seems possible that the technology could boost up the plant's revenue from selling the SNG and additional district heat. CO_2 capture and utilization together with the way of adding renewable energy share to the transport energy mix could also potentially affect the plant's image. Therefore, it is worth considering the technology as a good choice for retrofitting.

6. Conclusion

The study shows very high possibilities to implement the electrofuel technology with a waste-to-energy plant. Feasibility of the technology, however, is very sensitive with the cost of electricity and the SNG price. On the other hand, the investment cost plays just a minor role especially in larger capacities which economics of scale are greatly influential.

Furthermore, as the main source of income of waste-to-energy plant comes from district heat, it needs a careful consideration on the seasonal change in both district heat price and demand. Although the study indicates that the process with a higher capacity as of the winter case could be more profitable, the technical issue on lowering down the capacity during summertime is still questionable. Thus, some extra investment may be required in order to make such capacity decreasing possible.

Another critical parameter that could determine the feasibility of the process is electrolyzer efficiency since it has a very strong relationship to the electricity cost. An electrolyzer with higher performance is more preferable for the process even though it could mean to the greater capital cost. Efficient electrolysis technique such as SOEC could be a very promising alternative to be used instead of the conventional AEC.

Advantages of applying electrofuel technology with waste-to-energy plant is not limited to only utilize electricity for producing combustible gas, but it also promotes more income and good image to the plant as well as hints about another possible way to enhance national’s energy security which at the end could ultimately let the country to become energy import-independent.

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Appendix A: Input data

Operating hours

By using data from [29], different types of operating hours are determined for many purposes. The type is average operating hours which can be obtained by summing up the total monthly hours in each furnace line before averaging the sum. The result is as shown in Table A-1

Table A-1 Monthly average operating hour

Month	Average operating hours (hours)
Jan	702
Feb	640
Mar	672
Apr	720
May	600
Jun	398
Jul	539
Aug	532
Sep	541
Oct	744
Nov	705
Dec	690
Total	7483

The total hours is rounded up to 7500 hours. This represents a proper annual operating hours which is applied most of relevant calculation within this study. The monthly average hours are also used in some specific analysis such as variation in electricity and SNG production due to seasonal change.

Flue gas analysis

A starting point for all the calculations is to know amounts of flue gas and CO_2 emitted from the case study waste-to-energy plant. Composition of the flue gas is also crucial to select proper CO_2 capturing technology to be used in this study. The important data needed for this study are:

- Average concentration of CO_2
- Mass CO_2 emitted and trend
- Compositions of flue gas

Average concentration of CO_2

The useful concentration has to be in volumetric form. This can be done by dividing annual volumetric amount of CO_2 emitted with the annual flue gas emitted. To gain the volumetric amount of CO_2, data of average flue gas flow rate, its CO_2 content and operating hours of each month and line from [29] are used.

By multiplying the average flue gas flue rate with the operating hours. The amount of flue gas emitted can be obtained as in Table A-2. The volumetric CO_2 emitted can be gained in similar way but the CO_2 content has to be additionally multiplied. The value is represented as in Table A-3.

Table A-2 Monthly flue gas emitted in million Nm^3

Month	Line 1	Line 2	Line 3	Line 4	Total
Jan	47.8	77.5	70.9	46.4	242.6
Feb	50.2	58.8	70.1	41.1	220.3
Mar	55.5	74.8	58.0	33.9	222.2
Apr	53.7	70.3	71.0	38.7	233.8
May	52.4	69.4	74.3	11.9	208.2
Jun	33.2	66.3	12.7	20.2	132.4
Jul	10.5	75.5	47.2	43.7	176.9
Aug	32.2	7.0	75.1	47.1	161.4
Sep	35.6	33.0	66.1	36.1	170.9
Oct	54.4	77.6	75.5	46.8	254.3
Nov	52.8	75.2	71.3	45.9	245.2
Dec	48.8	78.3	75.9	34.5	237.4
				Annual emission	**2505.7**

Table A-3 Monthly CO_2 *emitted in million Nm*^{3}

Month	Line 1	Line 2	Line 3	Line 4	Total
Jan	5.1	8.3	7.4	5.3	26.0
Feb	5.3	5.9	7.3	4.8	23.3
Mar	5.8	8.2	5.9	4.0	23.9
Apr	5.4	7.6	7.2	4.6	24.8
May	5.2	7.5	7.5	1.4	21.7
Jun	3.3	7.1	1.3	2.0	13.7
Jul	1.0	8.1	4.5	5.1	18.7
Aug	2.5	0.6	7.8	5.3	16.2
Sep	3.5	2.6	6.9	3.6	16.7
Oct	5.8	8.1	7.7	5.3	27.0
Nov	5.3	7.6	6.9	5.3	25.1
Dec	4.9	8.0	7.8	3.3	24.0
				Annual emission	**261.0**

Finally, the average concentration is calculated based on these annual emission as 10.42%vol.

Mass CO_2 emitted

The total CO_2 emitted of each month obtained in the previous section is multiplied by density of CO_2 at the standard temperature and pressure (STP) of 1.977 kg/m^3 to gain the mass amount. From the result as in Table A-4, It can also be seen that a month with the lowest emission is June. This is then used as a basis for designing the capacity of Case B.

Table A-4 Mass CO_2 emission in ton

Month	Total (ton)
Jan	51.4
Feb	46.1
Mar	47.2
Apr	49.0
May	42.8
Jun	27.0
Jul	36.9
Aug	32.0
Sep	33.0
Oct	53.3
Nov	49.7
Dec	47.5
Annual emission	**516.0**

Compositions of flue gas

Main compositions of the plant's flue gas are as shown in Table 3.4. Amounts as of year 2013 is used for this study. To obtain concentration of each composition, known concentrations (i.e. concentration of CO_2 and O_2) are used together with the annual flue gas emission and density at STP of each component to gain the concentrations as %vol presented in Table A-5. The data is used for calculating flue gas molar flow rate needed to determine proper sizes of both VPSA and MEA.

Table A-5 Compositions of the flue gas

Composition	Concentration	
	%vol	ppm
NH_3	9×10^{-5}	0.9
CO_2	10.42	-
CO	2.3×10^{-3}	23.9
NO_x	3.4×10^{-3}	34.7
SO_x	1×10^{-4}	1.5
TOC	Negligible	
HCl	Negligible	
O_2	11	-
N_2	78.08	-

Appendix B: Heat potential

To calculate district heat potential of the Sabatier process, Aspen plus process modelling software is used. Assumptions (Table B-1) are made to represent a proper temperature range for heat exchanging between working fluid and district water streams. The process flow sheet and calculated heat duties are shown as Fig. B-1.

Table B-1 Assumptions made for process modelling

Inlet water (for hydrogen production) temperature (°C)	25
Outlet SNG temperature at the condenser (°C)	60

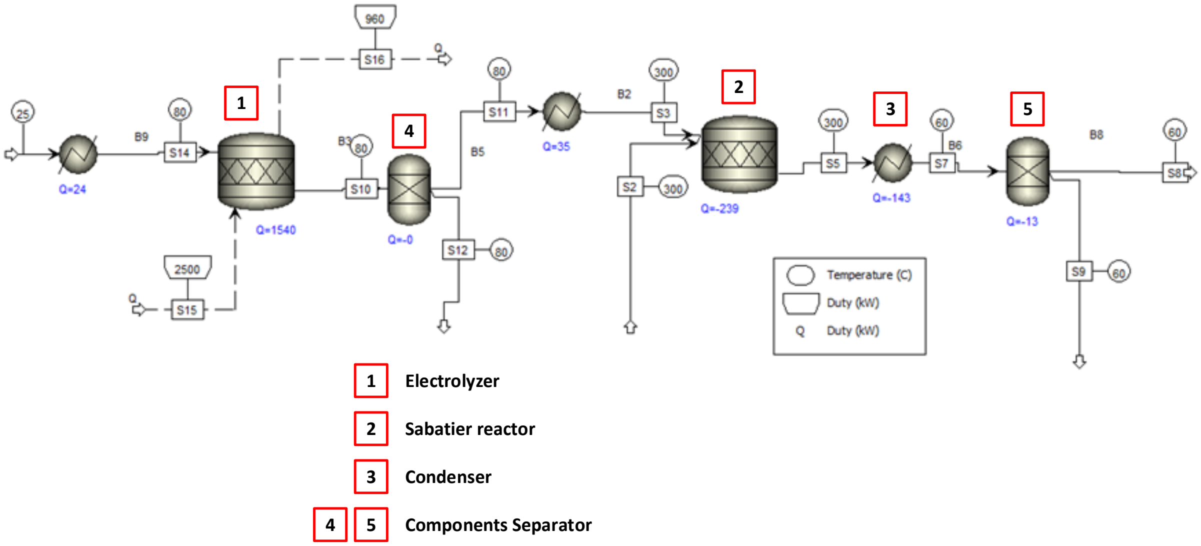
960
S16
25
80
B9
S14
Q=24
1
Q=1540
2500
S15
B3
80
S10
4
Q=-0
B5
S12
80
80
S11
Q=35
B2
300
S3
S2
300
2
Q=-239
300
S5
3
Q=-143
60
B6
S7
5
Q=-13
B8
60
S8
S9
60
Temperature (C)
Duty (kW)
Q Duty (kW)
1 Electrolyzer
2 Sabatier reactor
3 Condenser
4 5 Components Separator

Fig. B-1 Process diagram with main Sabatier process's equipment. Please note that the components separators used in the model are not actually existed. The equipment only represents way to separate components within the model.