CHALMERS
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Power-To-Gas Concepts Integrated with Biomethane Production through Gasification of Forest Residues

Economic Evaluation

Master's thesis in Sustainable Energy System Programme

PENPAKA KANNIKAPORN

Department of Space, Earth, and Environment
CHALMERS UNIVERSITY OF TECHNOLOGY
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MASTER’S THESIS 2018

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Supervisor: Johan Ahlström, Department of Space, Earth and Environment
Examiner: Stavros Papadokonstantakis, Department of Space, Earth and Environment

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Department of Space, Earth and Environment
Division of Energy Technology
Chalmers University of Technology
SE-412 96 Gothenburg
Telephone +46 31 772 1000

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ABSTRACT

Biomethane is an attractive transportation fuel due to an increase in demand of biofuels. Biomass gasification is a well-known technology used to produce biomethane from renewable feedstock, for instance, forest residues. The challenge with a high capital cost of gasification can be compromised by increasing the process efficiency, enhancing the yield of product biomethane, and operating in a larger plant scale. Power-to-gas technology is one of propitious solutions to help facilitating the biomass gasification by increasing the yield of biomethane while reducing the capital cost of CO_2 separation units and CO_2 emissions. To investigate the process integration possibility and the economic performances of biomass gasification integrated with power-to-gas technology, four different process designs, regarding the injection and ejection procedures of CO_2 and H_2, were evaluated in this study. The economic evaluation included total capital cost investment, production cost in terms of the levelized cost of fuel, and gross profit. The economic performances of different process designs, including the conventional base case for a stand-alone gasification plant, were compared, and the most preferable design was also suggested. The electrolysis technologies, alkaline and polymer electrolyte membrane (PEM) electrolysers, were compared in each configuration as well. According to the results, the process configurations integrated with alkaline electrolysis is relatively cheaper than PEM electrolysis, therefore, the process configurations integrated with alkaline electrolysis were only in focus. The overall economic evaluation shows that a process configuration, designed to feed H_2 so that a complete conversion of CO_2 can be achieved, is preferable due to its relatively low production cost and high profit. Although the configuration designed to increase process flexibility by recycling unreacted CO_2 back to the Sabatier inlet stream requires less total capital investment, the annual levelized cost of fuel of the former configuration is lower. By operating the former process configuration, the plant will potentially obtain 3 times more profit, compared to the conventional gasification, and the CO_2 emissions per unit of produced biomethane can also be minimized since a higher amount of biomethane (biofuel) can be obtained when using a similar quantity of utilized biomass.

Keywords: biomass gasification, biomethane, power-to-gas, Sabatier reactor, methanation, biomethane, SNG, cost estimation, economic evaluation, process integration.

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1. INTRODUCTION

1.1 Background

The interest of using biomethane as a renewable alternative to transportation fuel is increasing due to its advantages of storability and transportability within the existing gas distribution infrastructure [1]. Biomethane can be used for the same application as natural gas, for instance as a substitute for liquid transport fuels or a chemical feedstock, after conditioning the gas composition according to natural gas qualities on the market [2]. Two main biomethane production technologies, classified by the difference in feedstocks, are the biological (anaerobic fermentation) and the thermochemical (gasification) process [2]. The biochemical process is commonly used and well-developed with more than existing 200 operating plants in Europe, whereas the commercial biomethane production plant through gasification is firstly operated in Gothenburg while other plants are still under development in a demonstration stage [3].

The performance of the bio-chemical method highly depends on availability, composition, and degradability of feedstocks, leading to many drawbacks in terms of low biomethane yield, high retention time, and high investment cost regarding the process performance [4]. Due to a rising demand for biomethane consumption, only biological method is not enough to increase the market share in the transportation fuel. Therefore, other alternative methods and resources used to produce biofuel should be considered.

Consequently, biomethane production through gasification, using non-edible biomass such as forest residues (lignocellulosic biomass) as raw materials is a promising technology since it is more sustainable in terms of a capability of using abundant renewable feedstocks in a large-scale production considering an increase in biomethane demands, an overcoming of food production scarcity, and has relatively low negative environmental impact [5]. In addition, biomass gasification provides high performance in terms of cost and biomass-to-bioenergy efficiency, compared to other thermochemical processes such as pyrolysis and liquefaction [6]. However, the major challenge of biomethane production through gasification is the relatively high capital cost of gasification plants while the prices of natural gas are considerably low in today markets [7]. To address this challenge, a process efficiency and a yield of product biomethane together with a plant capacity should be increased to compete the relatively low prices of natural gas by decreasing a production cost per unit of produce biomethane.

Power-to-gas technology is one of propitious solutions that can be used to enhance the yield of biomethane produced through a biomass gasification plant. Power-to-gas concept is a process of converting electric energy into gas products which can be stored or used to produce biofuel. One example of a power-to-gas application integrated with gasification is a water electrolysis. Electrolysis entails that electricity is used to convert water into H_2 and O_2. The hydrogen gas is fed to react with the CO_2 in the product gas after the gasifier, which is otherwise captured and emitted to the atmosphere after biomethane purification stage. In this way, the yield of biomethane can be increased by the reaction between H_2 and CO_2, also known as Sabatier reaction [8], [9], while the capital cost of the CO_2 separation unit and CO_2 emissions per unit of produced biomethane can also be reduced. Moreover, the produced O_2 from electrolysis can be used as a gasifying agent when direct gasification technology is used.

With an increase in integration of intermittent electricity produced from renewable resources, a concept of power-to-gas is also applied to help balancing the fluctuating electricity grid by using an excess electricity (when electricity production is at peak) to produce H_2 through water electrolysis [2]. This application emphasizes the advantage of integrating power-to-gas technology with biomass gasification where H_2 is produced by electricity from renewable resources or when there is excess electricity available, and it can be kept in a H_2 storage to be used in many purposes (i.e. biomethane production).

Thereby, the integration of power-to-gas technology and the gasification from forest residues is an interesting option to improve the plant efficiency and enhance the yield of biomethane production, leading to a potential reduction in production expenses and increase in profit of the plant as well as a possibility of CO_2 emissions reduction.

In a previous study [10], process modelling of biomethane production through a direct oxygen blown biomass gasification plant was conducted in process simulation software Aspen Plus. The goal was to investigate the difference between the process performance of a conventional gasification plant and four process configurations integrated with power-to-gas technology, as the ways of CO_2 and H_2 injection and ejection were designed differently. However, the plant investment cost was not estimated. Therefore, the purpose of this study is to complete this previous work and fully evaluate the economic performance of a power-to-gas concepts integrated with biomethane production through gasification of forest residues, compared to a stand-alone gasification plant.

1.2 Aim

The aim of this project was to evaluate the production cost of biomass gasification integrated with different types of electrolysers based on the selected technologies and forecasted electricity prices. By achieving this aim, the suitable process design was suggested for the future works. The following tasks were performed:

- The cost estimation for overall production plant expense including total capital investment cost and annual product cost in terms of the levelized cost of fuel
- The scaling and overall costs including capital and production cost for electrolysers according to amount of hydrogen needed in the plant
- The sensitivity analysis of the cost of various process configurations with respect to electricity prices and capital investment cost of electrolysis technologies.

1.3 Limitations

The scope of this study was based on process modelling acquired from previous work by Gambardella and Yahya [10]. Four process configurations were compared based on the cost perspective. The electrolysis technology was focused on only alkaline and PEM electrolysers. The cost estimation may vary when considering different types of gasifiers, gasifying agents, and electrolysers, different production capacities, and geographic locations as well as difference financial factors (i.e. interest rate and the project lifetime). All these variations fall outside the scope of this study.

2. THEORY

In this chapter, a comprehensive description of the main process operation units is explained as a basis of understanding the overall process. The overall production plant included six main process areas: biomass preparation and drying, gasification, syngas cleaning, methanation, Sabatier process, and power-to-gas technology (electrolysis). The acronyms and more details in terms of process description and theory are also described in Appendix B and Appendix D, respectively.

2.1 Plant Configurations and Descriptions

The plant configuration is divided into two main different pathways: a conventional stand-alone biomass gasification plant and a biomass gasification plant integrated with power-to-gas technology. The main process operation units are similar for both process pathways, except that a Sabatier process and electrolysis are considered when the latter pathway is used.

2.1.1 Conventional Biomass Gasification – Base Case

The conventional biomass gasification in this study mainly consists of four process areas: biomass preparation and drying, gasification, syngas cleaning, and methanation process, as illustrated in Figure 4.1 [10] [22]. The process descriptions for each process area are demonstrated as following.

1) Biomass Preparation and Drying

The biomass is initially dried prior the gasification stage to reduce the moisture content from typically between 50 - 60% to below 15% of wet basis, since the heat demand for gasification can be significantly reduced with a lower initial moisture content of biomass feedstocks [11]. The typical energy content of dried biomass fuels is in the range of 15 - 20 MJ/kg dry ash free [12]. Since the moisture content of biomass can be reduced to 10 - 15% wet basis when a low-temperature drying is used [13], the belt dryer is, therefore, used in this study due to its low-temperature drying feasibility. Biomass is transferred to a sieve conveyor and transported along the dryer. The drying medium (i.e. hot air and steam) is blown through the conveyor and the biomass bed by fans [11].

2) Gasification

After the pre-treatment stage, the biomass is fed into a gasifier. Gasification is a thermochemical process converting any solid fuel (i.e. biomass) into gas products, in the presence of a gasifying agent such as air, oxygen, or steam. Gasification takes place at elevated temperatures between 500 and 1400 ºC, and at an atmospheric or high pressure up to 33 bars [5]. The produced gas mixture consists of CO, H_2, CO_2, H_2O, CH_4, and small quantities of hydrocarbons, tars, impurities, as well as N_2 when air is used as a gasifying agent in the process [14]. In this study, a direct oxygen-blown gasifier is used.

3) Syngas Cleaning

Syngas cleaning (or syngas conditioning) is required to eliminate some impurities containing in the gas mixture that could affect the downstream process or the product gas utilisation. The gas treatment also helps remove undesired gas compounds and adjust the gas component ratio to a proper value according to the downstream process requirements [15]. The syngas cleaning process, considered in this study, includes particulates separation by a ceramic filter, tars removal using a catalytic reforming, steam and moisture separation, and acidic gases removal using rectisol unit for an upstream process.

4) Water-Gas-Shift and Methanation Process

The produced gas mixture after gasification consists mainly of carbon monoxide, hydrogen, and methane [16]. Since the gas mixture contains an uneconomically low quantity of methane, the amount of methane is normally increased by using the technique called methanation process. The composition ratio of H_2 and CO in the gas mixture has to be adjusted through a water-gas-shift reaction in order to obtain the desired gas ratio of H_2/CO prior to a methanation process [17]. Methanation is a primary technique used to produce CH_4 and H_2O from the reaction between CO and H_2 [18].

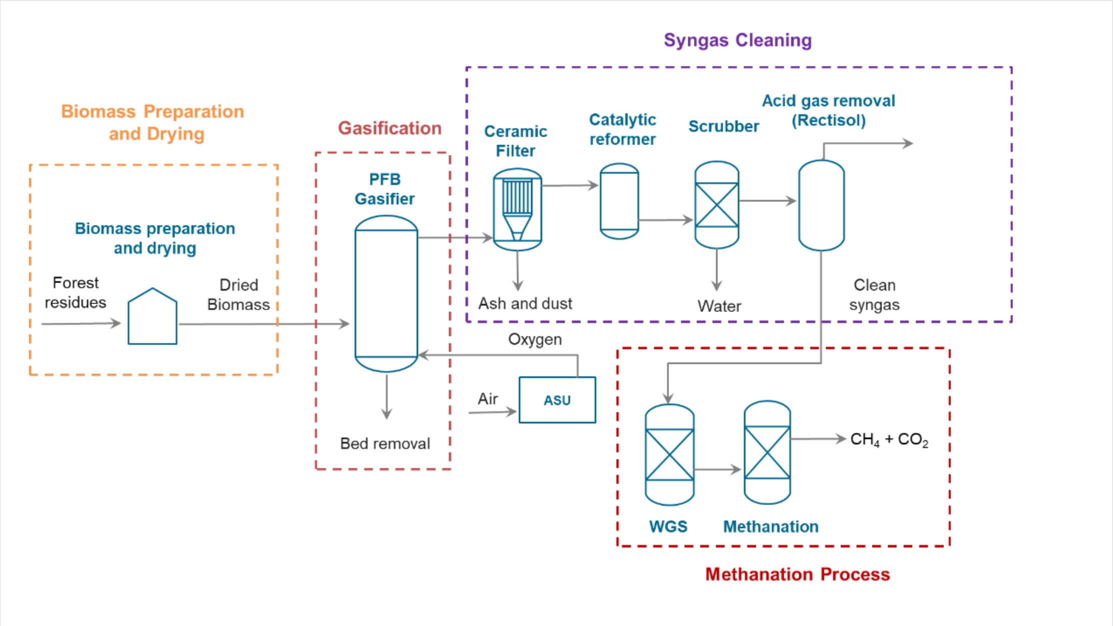

Figure 2.1 The process scheme of biomass gasification – Base case [10] [22]

After the methanation process, the gas mixture contains mostly biomethane and carbon dioxide as well as low amounts of steam. To obtain a market purity standard (Wobbe Index) of biomethane, CO_2 in a gas mixture can be handled in two ways: be separated and released to the atmosphere or to be injected with H_2 to Sabatier process in order to enhance the yield of product methane. For a conventional gasification plant, the former alternative is normally used.

2.1.2 Biomass Gasification Integrated with Power-to-Gas Technology

Since the typical upstream processes for gasification are the same as using in a conventional pathway, the focus is only on the Sabatier process in this section. As previously mentioned, the gas mixture after methanation contains mostly CH_4 and CO_2 (produced mainly in water-gas-shift reaction) [10]. In the Sabatier process, CO_2 in a gas mixture is proceeded to react with an additional H_2 in Sabatier process to enhance the yield of product biomethane. The Sabatier process is an exothermic reaction where CH_4 and H_2O are formed by the reaction between CO_2 and H_2 in the presence of catalyst. In this study, the Sabatier reactor designed by Turbo SE and MAN Diesel [19] was used. By operating a Sabatier process, not only the yield of biomethane can be increased, but also the need for carbon dioxide separation and the amount of CO_2 emission can be decreased. However, some additional amount of pure hydrogen gas is required in the process, which is considered as an operating cost penalty.

The use of power-to-gas technology in terms of water electrolysis integrated with biomass gasification is one example that can compromise the high cost of hydrogen required in a Sabatier process. The electrolysis systems used in this study are alkaline electrolysers (liquid electrolyte) and polymer electrolyte membrane (PEM) electrolysers since they are the most mature and recently available in the market. The electricity that used to convert water into H_2 and O_2 could come from the intermittent renewable energy sources, such as wind and solar energy, but more research is required to ensure the utilisation of this energy while maintaining the stability of the grid. By applying the power-to-gas technology integrated with the gasification process, it can not only enhance the yield of product biomethane, but it can also benefit the plant when the produced O_2 gas from electrolysis can be used as a gasifying agent for a direct gasification.

Due to a potential benefit of integrating a power-to-gas concept with a biomass gasification, it is interesting to investigate the technical and economic performance of the overall process. In previous study [10], Gambardella and Yahya conducted the process modelling and evaluate the technical process performance of the gasification integrated with a power-to-gas concept. Four different process designs regarding the injection and ejection of H_2 and CO_2 into a Sabatier process was modelled and investigated. However, the investment cost was not done in a previous study. In this study, it is therefore of interest to continue the work by evaluating the economic performance of these four process configurations together with a conventional stand-alone gasification process and thus to suggest a preferable configuration regarding process requirements and financial parameters.

The upstream process for all process configurations is the same, as well as all operating parameters used in the process modelling (i.e. pressure, temperature, catalyst type and loading). The difference between each process configuration is the injection and ejection procedures of CO_2 and H_2. The description of four process modelling schemes are demonstrated as following.

1) Configuration 1

The produced H_2 from electrolysis is pressurized and sent to Sabatier reactor together with a product gas mixture from the methanation process and some amount of unconverted CO_2 recycled from CO_2 capture units (Figure 2.4). In this case, the plant operational flexibility can be increased when the unreacted CO_2 is recycled back to the Sabatier inlet stream, which may have a significant effect on the overall performance.

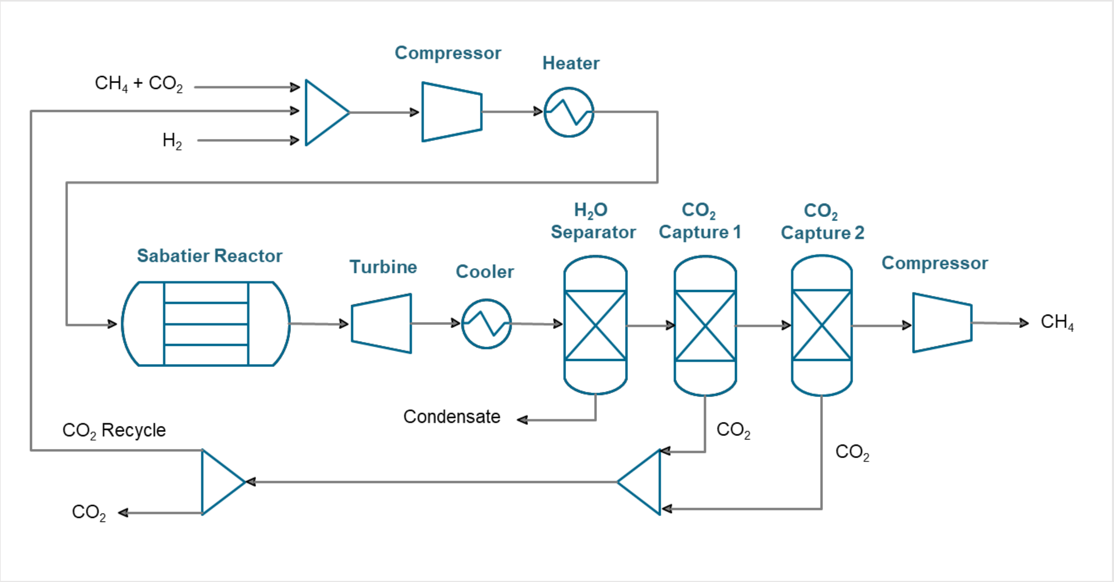

Figure 2.2 The process scheme of the Sabatier process for Configuration 1 [10]

2) Configuration 2

The produced H_2 from electrolysis is pressurized and sent to the Sabatier reactor together with a product gas mixture from methanation process. For this configuration, the H_2 gas is fed to achieve a complete conversion of CO_2 (Figure 2.5). Consequently, there is no CO_2 separator in the process and the CO_2 emission is minimised. Nevertheless, an excess amount of H_2 is needed in order to achieve a complete conversion of CO_2, therefore the H_2 removal unit is still required in the downstream process. In this case, the process operation range is limited by an abundant availability of H_2, thus decreasing the process flexibility. For instance, it can be a case when there is no an excess amount of H_2 in a Sabatier reactor, leading to a difficulty to achieve a complete conversion of CO_2.

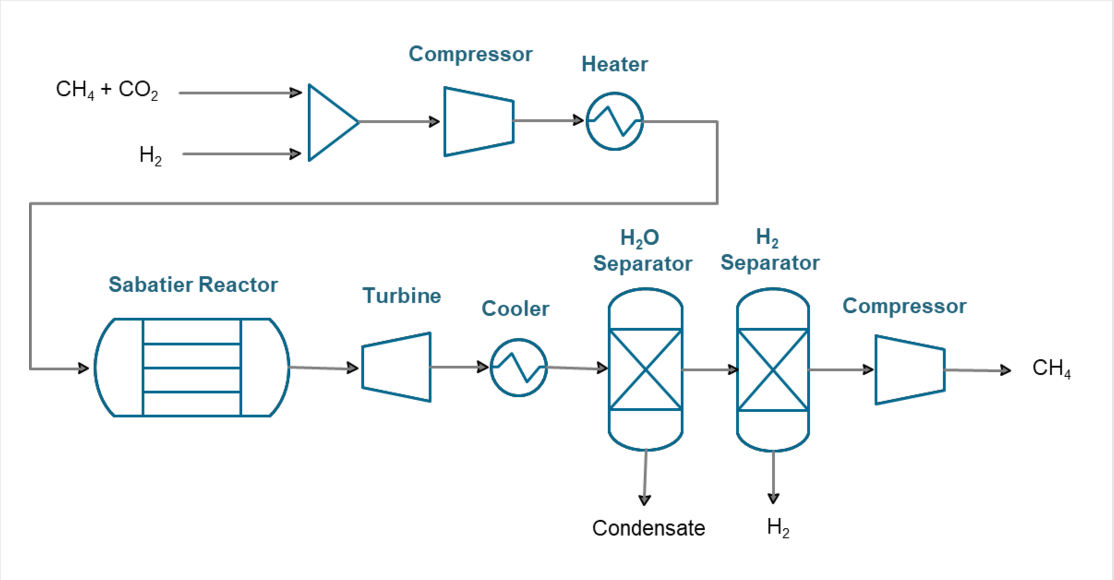

Figure 2.3 The process scheme of the Sabatier process for Configuration 2 [10]

3) Configuration 3

For this configuration, only a desired amount of CO_2 is sent to a Sabatier reactor (Figure 2.6). It provides an operational flexibility and a better optimization of the reactor size. The CO_2 from both the methanation and the Sabatier processes is removed in the same CO_2 capture units. In addition, the H_2 removal unit is still required to achieve the maximum operational flexibility since the process can be adjusted in a wider operational range regarding a CO_2 and H_2 availability and process requirement, as well as plant capacity.

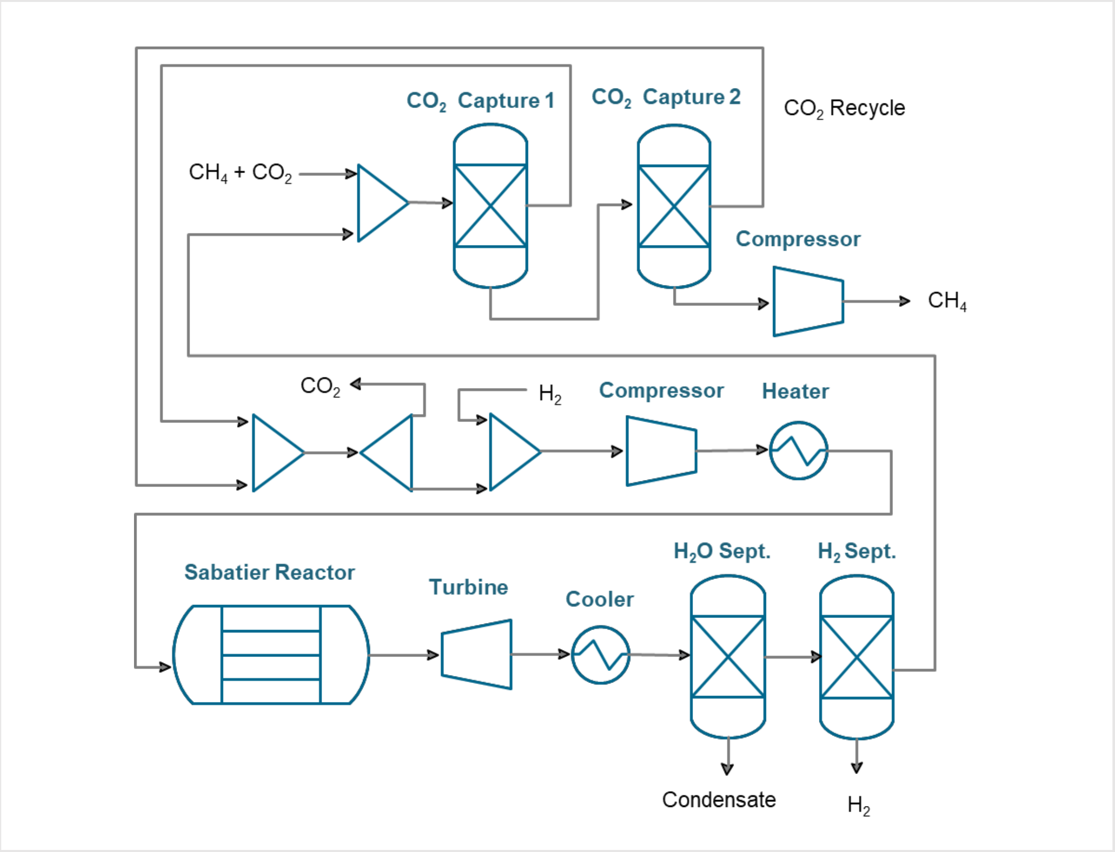

Figure 2.4 The process scheme of the Sabatier process for Configuration 3 [10]

4) Configuration 4

In this case, the gas stream from the methanation process is mixed with the dried product stream of the Sabatier process. The CO_2 is removed from the product stream prior to the Sabatier process. This configuration is identical to Configuration 3, but no H_2 removal is required in Configuration 4 (Figure 2.7).

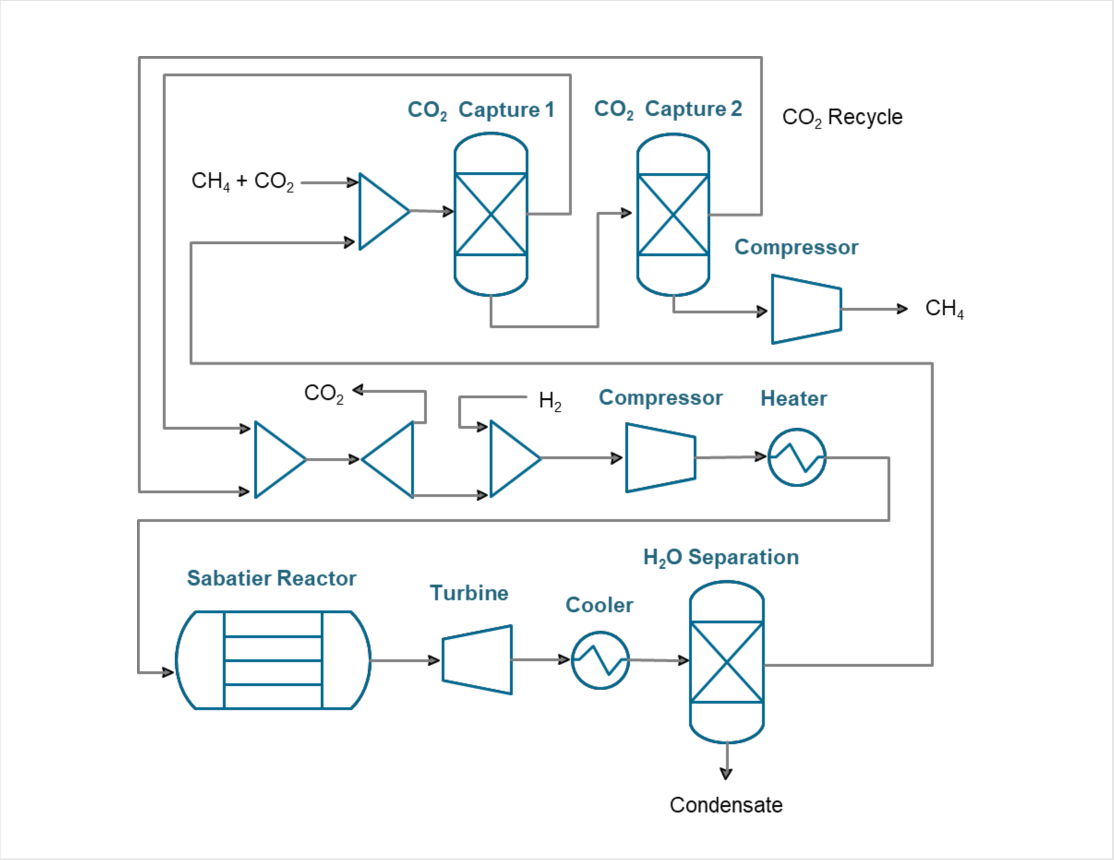

Figure 2.5 The process scheme of the Sabatier process for Configuration 4 [10]

To summarize, the difference between process configurations can be categorised into two aspects: an operational flexibility, and a requirement of CO_2 and H_2 separation units.

The difference between Configuration 1 and 2 is that a CO_2 capture unit is required, but a H_2 removal unit is not needed for Configuration 1 and vice versa for Configuration 2. In Configuration 1, some amount of unreacted CO_2 can be adjusted and sent back to a Sabatier process. In this case, the operational flexibility for Configuration 1 is relatively higher than for Configuration 2, which has no operational flexibility regarding an operational range and plant capacity, as explained in Configuration 2.

For Configuration 3 and 4, a higher operational flexibility can be obtained, compared to Configurations 1 and 2, since only a desired amount of CO_2 is sent to a Sabatier reactor after separating out of the outlet streams from both methanation and the Sabatier processes. The Configuration 3 and 4 are identical except only that H_2 separation unit is not required for Configuration 4 in order to investigate the economic performance when the operational flexibility is reduced regarding an ability to remove H_2.

3. METHODOLOGY

In this study, the project work is divided into two parts: the economic analysis in terms of an overall investment and production cost as well as a sensitivity analysis of process configuration based on the forecasted electricity prices and electrolysis system costs. The financial factors used in this study are assumed as 20 years of project lifetime and 10% interest rate. All results evaluated in this study is calculated on the basis of the Euro currency rate at the year of 2017. The financial parameters and cost variables used in the cost estimation can be found in Appendix A.

3.1 Economic Analysis

To evaluate the economic performance, total capital investment (TCI), annual production cost and annual gross profit are used as economic performance indicators. The detail cost estimation in terms of a plant location consideration, labour cost, depreciation cost, and income tax, is not considered in the economic evaluation of this study. The operating factors for electricity prices were based on a Swedish context; the other financial parameters and cost variables used in the cost estimation were retrieved from reference literature, as shown in Appendix A.

3.1.1 Total Capital Investment

The total capital investment of the process equipment is estimated in two steps. First, most of the installed equipment costs are estimated based on previous literatures using similar technology, feedstock, and operating parameters as the process modelling conducted by Gambardella and Yahya [10]. Second, the capital costs of some conventional equipment units, e.g. the carbon dioxide capture unit and the heat exchanger network are estimated by using the cost estimation tool Aspen In-plant cost estimator followed by scaling methods according to a defined plant capacity.

To investigate the difference in total capital investment cost for each plant capacity, the individual installed equipment cost is scaled-up using the power relationship known as the six-tenths factor rule [21], as shown in Equation (3.1).

$$C = C_o \times \left(\frac{S}{S_o}\right)^f \qquad (3.1)$$

The current cost values in 2017 million euros (C) is obtained by using the cost of a reference installed equipment (C_o) of the reference equipment capacity (S_o) The specified equipment capacity (S) is scaled-up by using the cost scaling exponent (f) which is different depending on equipment types and maturity of equipment technologies.

The reference installed equipment costs, unit capacities, scaling factors, and other additional cost factor used to evaluate total capital investment are shown as in Appendix C.

3.1.2 Production Cost

The production cost was considered in terms of the levelized cost of fuel (LCOF), which is the annual production cost divided by the annual fuel output (M). The annual production cost can be calculated by the summation of biomass feedstock cost (F), electricity cost (E), annual capital charge (C_p) and operating and maintenance cost (O), subtracting the annual by-product revenue (R), which consists of O_2 and steam revenues. The levelized cost of fuel is calculated by using Equation (3.2) [22].

$$\mathrm{LCOF} = \frac{F + E + C_p + O - R}{M} \qquad (3.2)$$

The annual capital charge (C_p) and annuity factor (r), for an assumed interest rate (i) and a plant lifetime (n) are calculated as shown in Equation (3.3) [23] and Equation (3.4) [21], respectively.

$$C_p = \frac{r\,(\mathrm{NPV})}{1 - (1 + r)^{-n}} \qquad (3.3)$$

$$r = \frac{i\,(1 + i)^{n}}{(1 + i)^{n} - 1} \qquad (3.4)$$

3.1.3 Gross profit

In this study, the gross profit was calculated without a consideration in depreciation and income taxes. The gross profit is determined by subtracting the total production cost out of the total product sale revenues, as shown in Equation (3.3).

$$\text{Gross Profit} = (M_c + R) - (F + E + C_p + O) \qquad (3.3)$$

The total product sale revenues include the revenue from selling product methane (M_c), and revenue from selling by product (R).

3.2 Sensitivity Analysis

In the second part, sensitivity analysis of process configurations is conducted in a computational program like Microsoft Excel to estimate the cost of process configuration regarding the fluctuation of forecasted electricity prices and the predicted electrolysis system capital investment.

Since the electricity price is a major factor influencing the production cost, it is interesting to investigate the deviation of the plant production cost regarding the variation of electricity prices. The base electricity price is 35 €/MWh as retrieved [25]. The electricity price is varied in the range between 15 - 55 €/MWh in a sensitivity analysis and the production cost in terms of the levelized cost of fuel for each process configuration is compared.

The reduction of electrolysis system price regarding the year of the plant investment is also considered to compare the possible capital cost reduction of electrolysis system when the plant is invested in 2030, instead of 2017.

4. RESULTS AND DISCUSSION

In this chapter, the economic results of the base case and all process configurations are presented and compared in order to determine whether the biomass gasification plant should be integrated with a power-to-gas technology and to identify which process configuration is preferable to operate regarding technical and economic performance.

4.1 Economic Performance

The economic performance indicators used in this study includes the total capital investment, the annual production cost and the annual gross profit.

4.1.1 Total Capital Investment

The total capital investment (TCI) of all configurations including the base case for the plant capacities of 20 - 340 $MW_{LHV,\ biomass}$ is demonstrated in Figure 4.1. The figure shows that the total capital investment per biomass thermal input of all cases decrease with an increase in the plant capacity, regarding the economy of scale. The figure shows that the plant integrated with a PEM electrolysis requires a higher capital investment cost compared to alkaline electrolysis (A) for all configurations, based on the fact that a cost of PEM electrolysis is more expensive than that of an alkaline. Therefore, the alkaline electrolysis system is selected and only focused on in the following sections. Apart from the base case, the graph shows that Configuration 1A reaches the lowest capital investment cost.

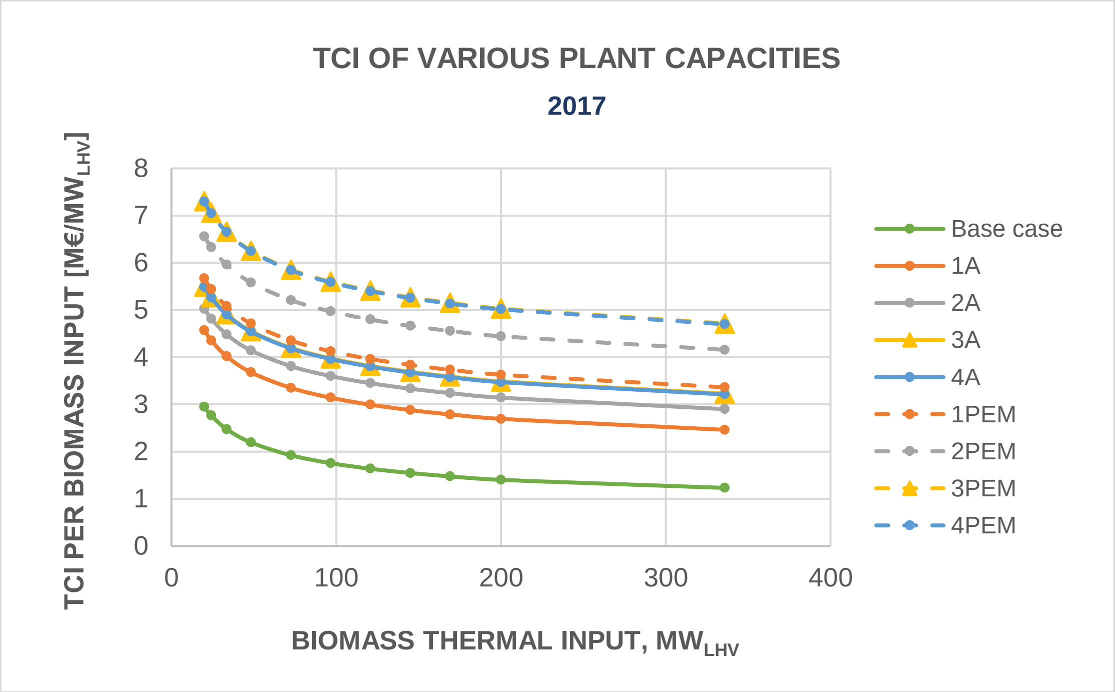

Figure 4.1 The total capital investment of the different process configurations

From Figure 4.1, the result shows that a plant capacity of 200 $MW_{LHV, biomass}$ should be selected since the total capital investment tends to be stable and does not noticeably decrease with a further increase in the plant capacity. As a reason, the plant capacity of 200 $MW_{LHV, biomass}$ is, therefore focused in the following results in order to investigate the benefit of operating the plant at this capacity. The result of total capital investments of all configurations for 200 $MW_{LHV, biomass}$ is illustrated in Figure 4.2.

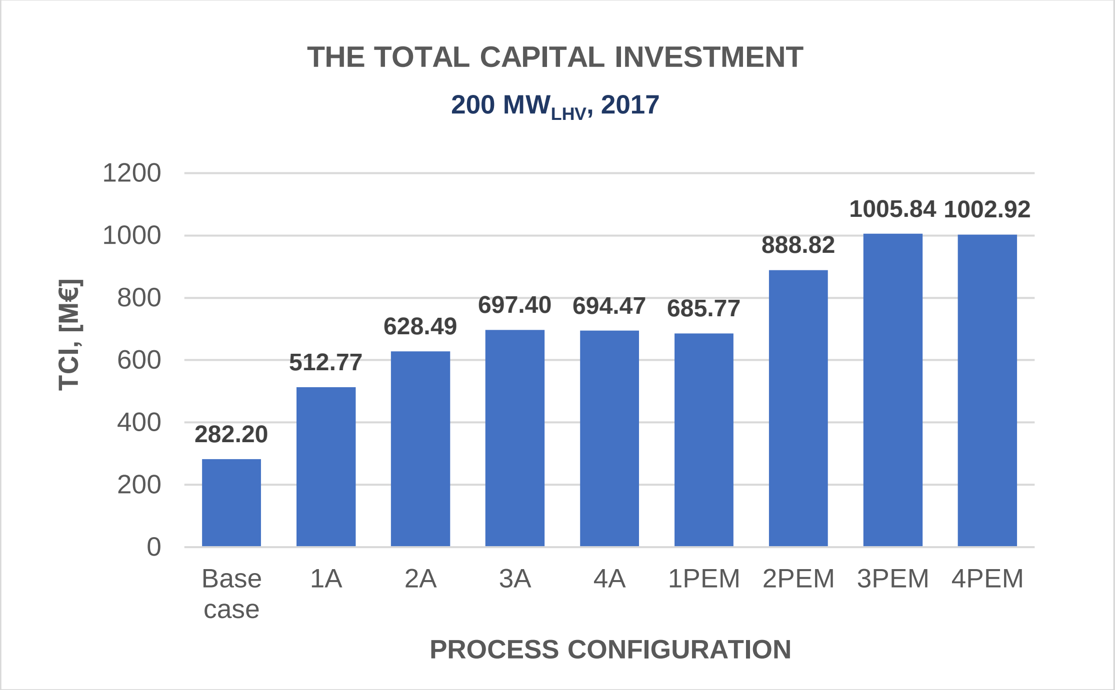

Figure 4.2 The total capital investment of the different process configurations

To demonstrate the capital investment for each process equipment unit, the cost fractions of total capital investment for base case and four process configurations, described in the chapter 3, are presented in Figure 4.3 and Figure 4.4, respectively.

For the base case, the main investment costs come from the syngas cleaning, the heat exchanger network, and the methanation. The capital investment is significantly increased when the process is integrated with the electrolysis system. Consequently, the electrolysis system becomes the major cost of the overall production plant. From Figure 4.4, the cost of other parts of process equipment for all configurations are relatively the same, except for the electrolysis part, leading to a great difference in capital cost for each configuration. Note that, the unscheduled equipment cost is included as a faction of an approximately 10 % of total capital investment. It is generally considered as an indirect cost for miscellaneous equipment and spare parts, which may lead to an additional cost in the project.

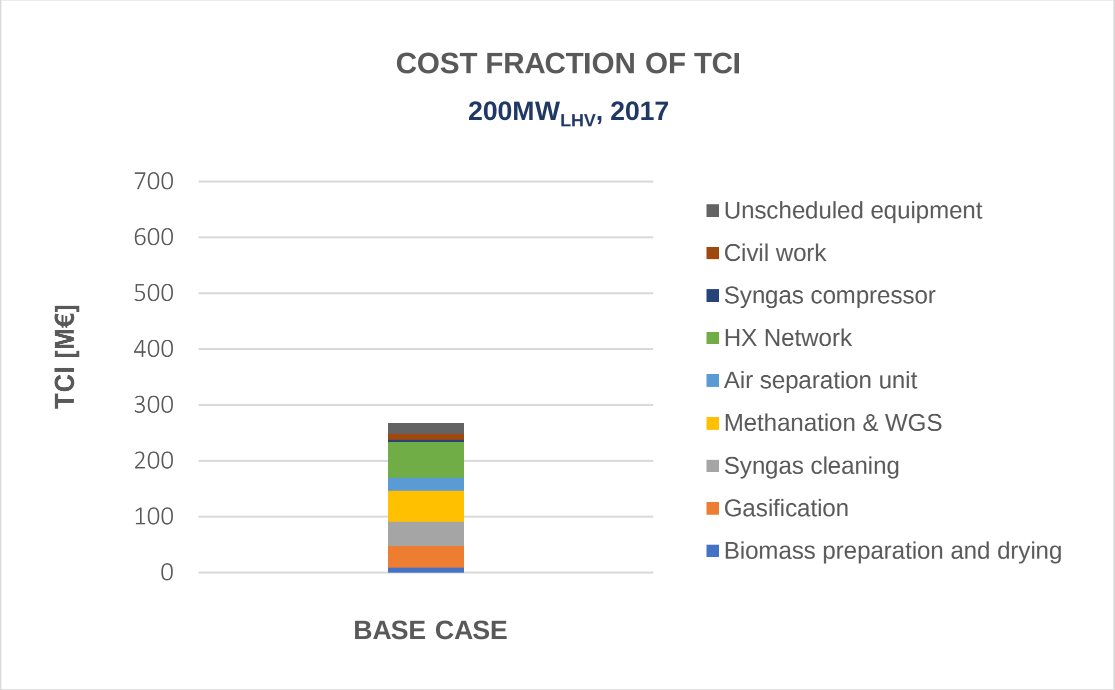

Figure 4.3 The cost fraction of total capital investment for base case

The main reason that Configuration 1A required the lowest capital investment is that it requires less additional hydrogen to the process and no hydrogen removal unit is installed, therefore the capital equipment cost of electrolysis system decreases. For Configuration 2A, the carbon dioxide capture unit is not installed in the plant, leading to a less capital investment compared to Configuration 3A and 4A. The Configuration 3A and 4A are identical except that the hydrogen removal unit is installed in Configuration 3A. This is a reason that the difference between the two of them can be hardly noticed.

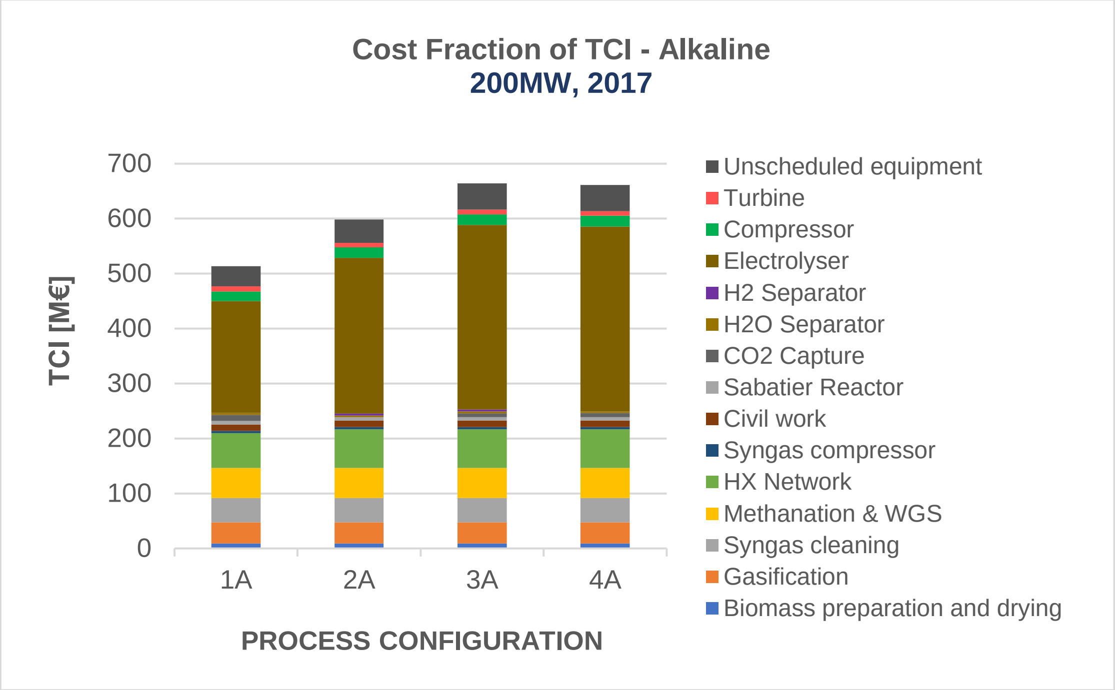

Figure 4.4 The cost fraction of total capital investment for process configurations integrated with an alkaline electrolysis system for 2017

4.1.2 Production Cost

The production cost is evaluated in terms of levelized cost of fuel (LCOF). It represents the minimum cost required to produce a certain amount of fuel, in this case, it also indicates the annual breakeven price of fuel production for each process configuration under specific technical and economic circumstances. The figure shows that the levelized cost of fuel decreases with an increase in plant capacity since the total capital cost and production cost used to produce a unit of methane reduce. The result shows that process Configuration 2A requires the lowest levelized cost of fuel, followed by Configuration 1A, 4A, 3A, and the conventional gasification (base case), as shown in Figure 4.5.

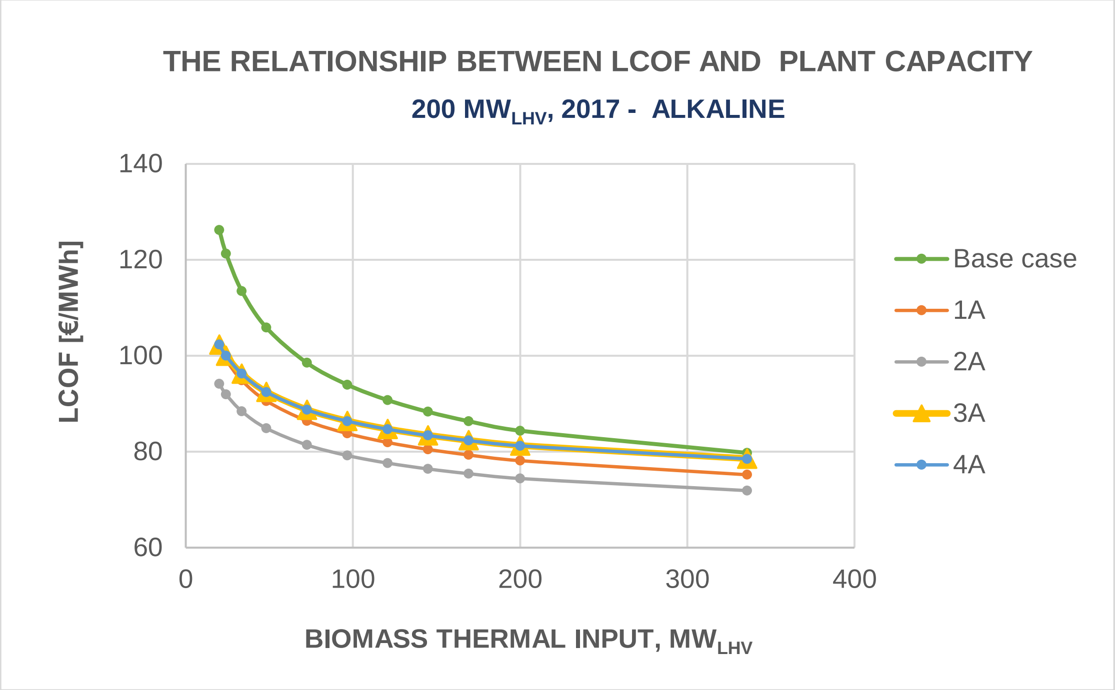

Figure 4.5 The relationship between the levelized cost of fuel (LCOF) and plant capacity base on biomass thermal input.

The reason is that the yields of produced methane for all configurations are significantly increased, approximately twice as much compared to the base case, when the product gas is further processed in the Sabatier reactor. Therefore, the production cost per unit of product methane for all process configuration are relatively lower than for the conventional one.

The levelized cost of fuel for Configuration 2A is lower than for Configuration 1A, in contrast with the total capital investment as shown in the previous section. The reason is that a higher yield of methane can be produced in Configuration 2A compared to Configuration 1A, and that it is high enough to overcome the difference in annual capital charge as well as the operating and maintenance cost. Although the higher yield of methane can also be obtained by operating Configuration 3A and 4A, the revenues from produced methane are not high enough to cover their significantly higher capital cost. As a result, the amount of methane production is the main factor influencing the production cost in terms of levelized cost of fuel. Thus, Configuration 2A is a preferable process design that requires the least annual expenses. The relationship between the total capital investment and the levelized cost of fuel for 200 MWLHV, biomass is demonstrated in Figure 4.6.

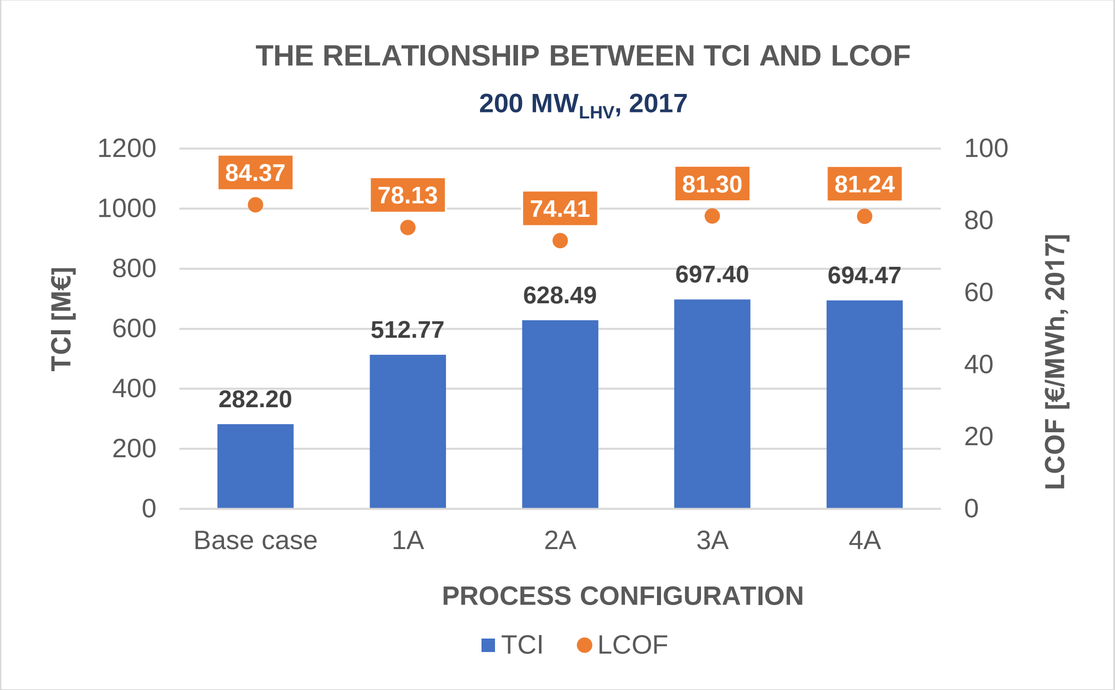

Figure 4.6 The relationship between total capital investment and levelized cost of fuel for 200 MW$_{LHV,\ biomass}$ plant capacity

The distributions of production cost factors are also illustrated, see Figure 4.7, to identify the cost components that highly affected to the levelized cost of fuel. The figure shows that the electricity cost is the main contribution to the total production cost, followed by the capital charge due to the capital investment in the electrolysis system. From Figure 4.7, the revenues from biomethane and by-products are also compared to the overall production cost. It shows that Configuration 2A requires less electricity consumption to produce a relatively equal amount of biomethane as produced from Configuration 3A and 4A. Consequently, the levelized cost of fuel for Configuration 2A is the lowest one.

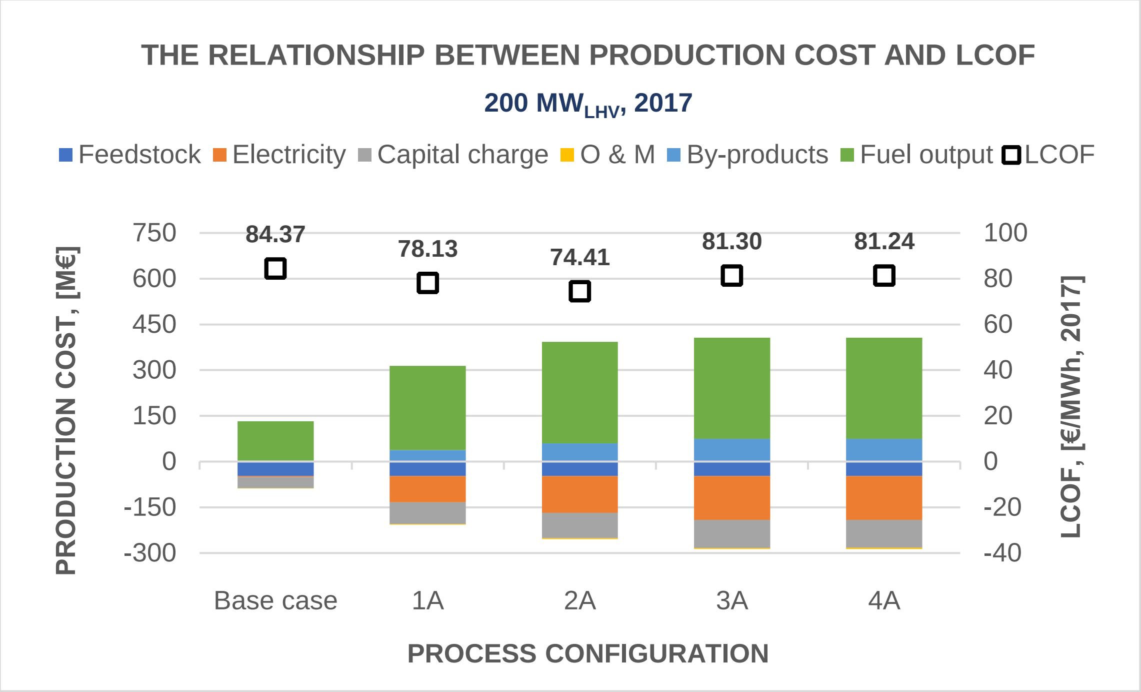

Figure 4.7 The relationship between total capital investment, production cost, and levelized cost of fuel for 200 MW$_{LHV,\ biomass}$ plant capacity

4.1.3 Gross Profit

The gross profit is evaluated to determine the possible earning of each process configuration, as shown in Figure 4.8. The location of the plant, the labour rate, depreciation, and the tax income are not considered. The result shows that Configuration 2A is preferable as it can provide a higher profit compared to other configurations, similarly to the previous section. Nevertheless, the critical drawback of Configuration 2A is that the operational flexibility is imposed since the Sabatier reactor size and amount of H_2 feed are limited by the plant capacity. Therefore, it is essential to take a trade-off between the highest possible profit and the process operational flexibility of the plant in to consideration when deciding the suitable plant configuration.

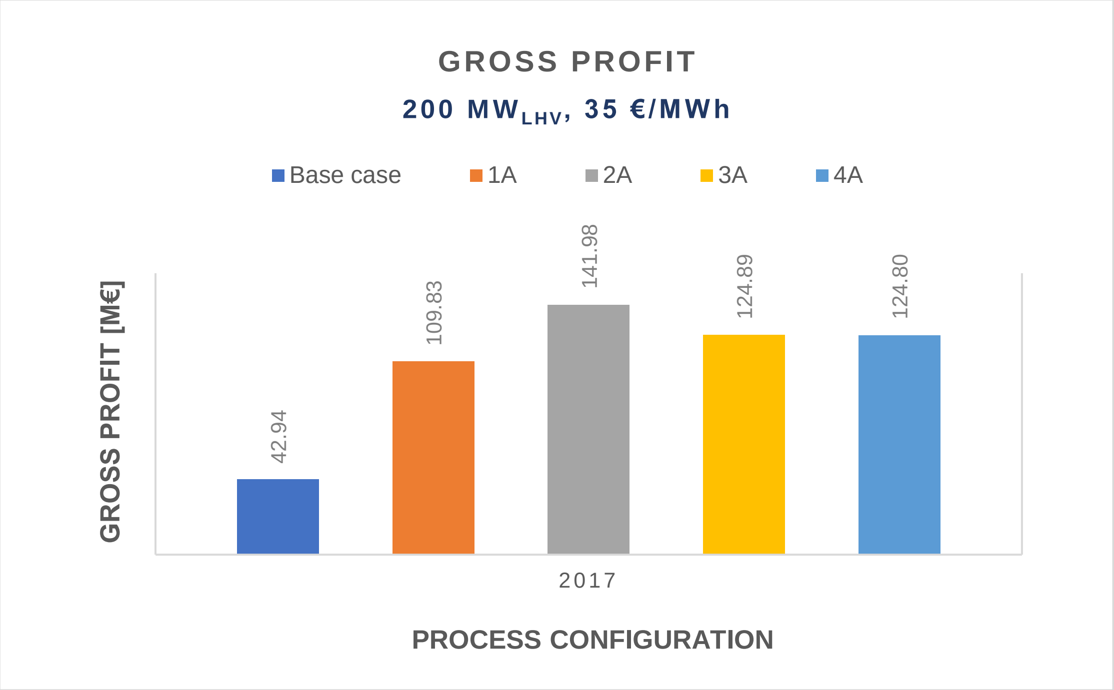

Figure 4.8 The gross profit of all process configurations for 200 $MW_{LHV,\ biomass}$ plant capacity

4.2 Sensitivity Analysis

The sensitivity analysis is divided into two aspects: the effect of electricity prices and the effect of the predicted electrolysis system prices. This analysis is evaluated to investigate the changing trend and the suitable conditions for each configuration regarding the process variables.

4.2.1 Sensitivity Analysis on Electricity Prices

The relationship between the levelized cost of fuel and electricity prices for all configurations is illustrated in Figure 4.9. The figure shows that Configuration 2A is still the preferable process design for the electricity price up to 55 €/MWh while it tends to be more preferable for Configuration 1A if the electricity price is higher than 55 €/MWh, which can compete the Configuration 2A. For Configuration 3A and 4A, it is better to operate when the electricity price is below 30 €/MWh comparing to Configuration 1A, and at electricity below 15 €/MWh comparing to Configuration 2A. The reason is that more biomethane can be produced in Configuration 3A and 4A compared to Configuration 1A and the revenues from biomethane and by-products can compensate the electricity cost with the lower electricity unit price used in the process. Anyhow, all four configurations should not be operated when the electricity is higher than 45 €/MWh, since the base case requires the least levelized cost of fuel of all cases.

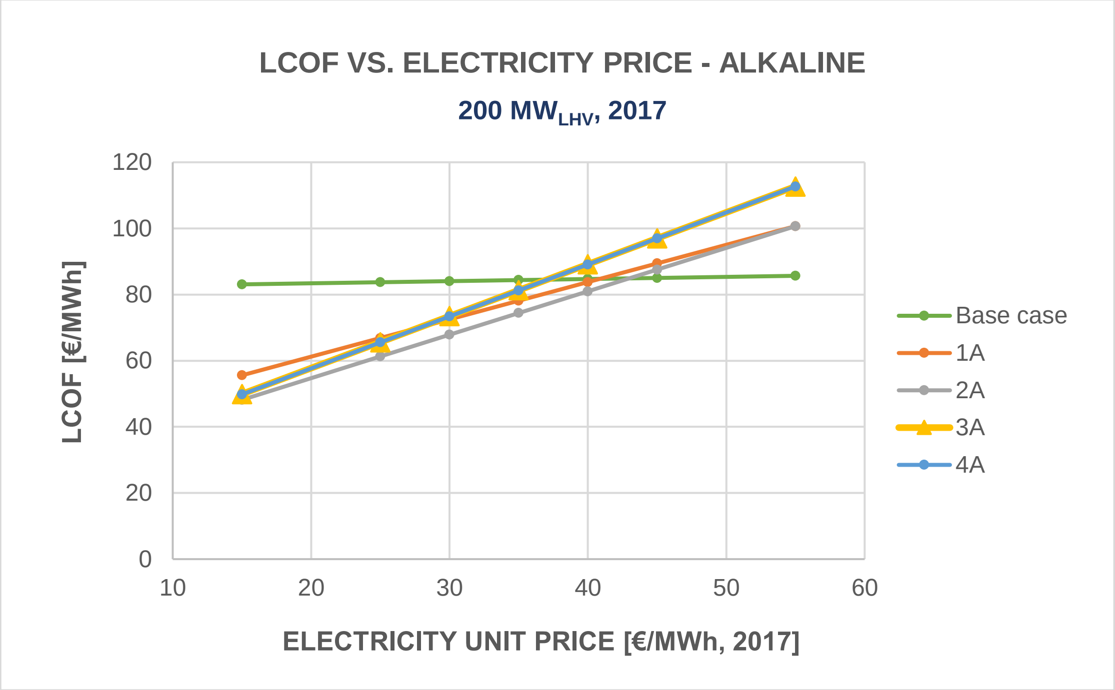

Figure 4.9 The relationship between the levelized cost of fuel and the electricity prices

In contrast with the levelised cost of fuel, the result of a graph illustrating the relationship between gross profit and electricity prices, see Figure 4.10, shows that all process configurations integrated with power-to-gas technology is more profitable than the conventional base-case gasification when it is operating at an electricity price below 55 €/MWh, instead of 45 €/MWh as shown in Figure 4.9. Moreover, the base case configuration tends to obtain more profit than the Configuration 3A and 4A when operating at the electricity price higher than 55 €/MWh.

In addition, although the sensitivity analysis for the electricity prices above 55 €/MWh is not calculated, the figure shows the trend of a decreasing profit regarding an increase in electricity prices, thus, it is more reasonable to operate only a base case configuration rather than considering investing the power-to-gas technology when the electricity price in the market is higher than 70 €/MWh.

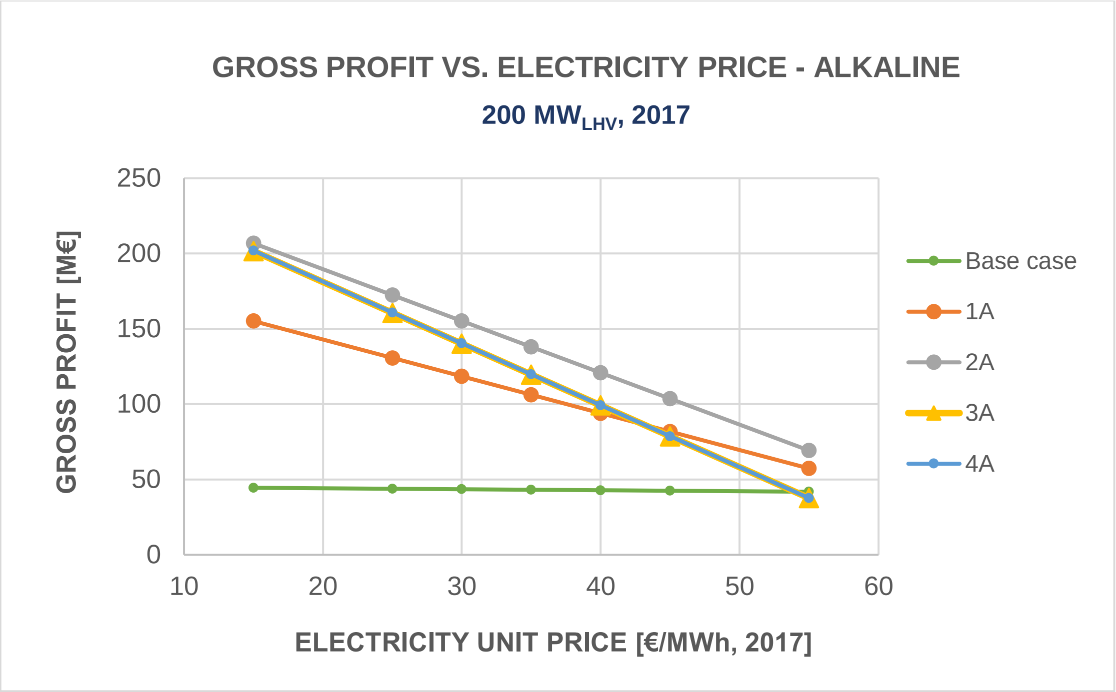

Figure 4.10 The relationship between gross profit and predicted electricity prices.

In comparison between Figure 4.9 and Figure 4.10, the results show that although the levelized cost of fuel for the Configuration 2A is higher than the base case when the electricity price is higher than 45 €/MWh, the relatively higher profit can still be obtained for Configuration 2A. The reason is that Configuration 2A produces significantly higher amounts of methane comparing to the base case configuration, and that can cover the higher levelized cost of fuel for the overall plant.

Anyhow, it should be noted that the prices of feedstocks, biomethane, by-products, and other costs influencing in total production cost are assumed to be fixed and do not change while the electricity prices are varied and may drop in the future scenarios, for a simplification of economic evaluation.

4.2.2 Sensitivity Analysis on Electrolysis System Cost

According to a study of Bertuccioli et al. [24], see Appendix D, the capital cost for the electrolysis system is expected to be significantly decreased in the future. Therefore, the total capital investments for all process configurations are estimated based on the predicted electrolysis system cost for the year of 2030. The total capital investment and the cost fraction for the plant invested in 2030 are illustrated in Figure 4.11. The figure shows that the total capital investment is decreased at approximately 100 million euros when the plant is invested in 2030, instead of 2017, as the technology matures. It also shows that the electrolysis system cost distributes in a smaller proportion to the total capital investment compared to other equipment cost, leading to an increase in a competitiveness of the technology.

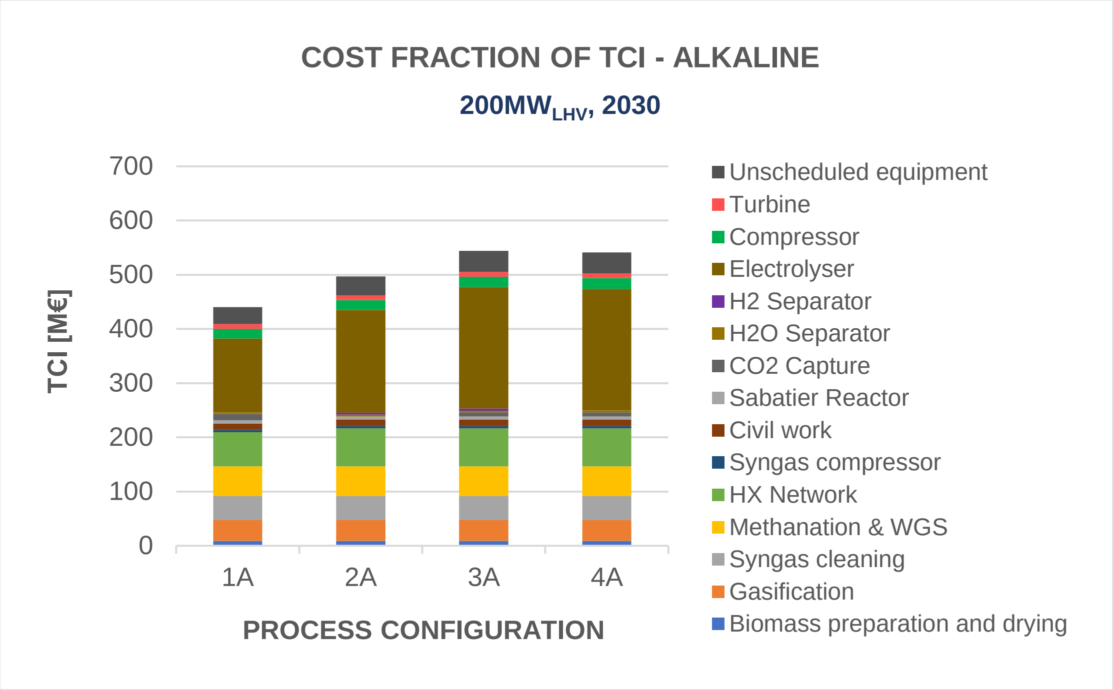

Figure 4.11 The cost fraction in total capital investment for process configurations integrated with an alkaline electrolysis system for 2030

The gross profit for all configurations, except the base case, will increase approximately 8 - 12% when the plant will be invested at the year of 2030, instead of 2017, as shown in Figure 4.12. The change gross profit between 2020 and 2030 was slightly different due to a little deviation of the electrolysis system capital cost, as mentioned in previous section.

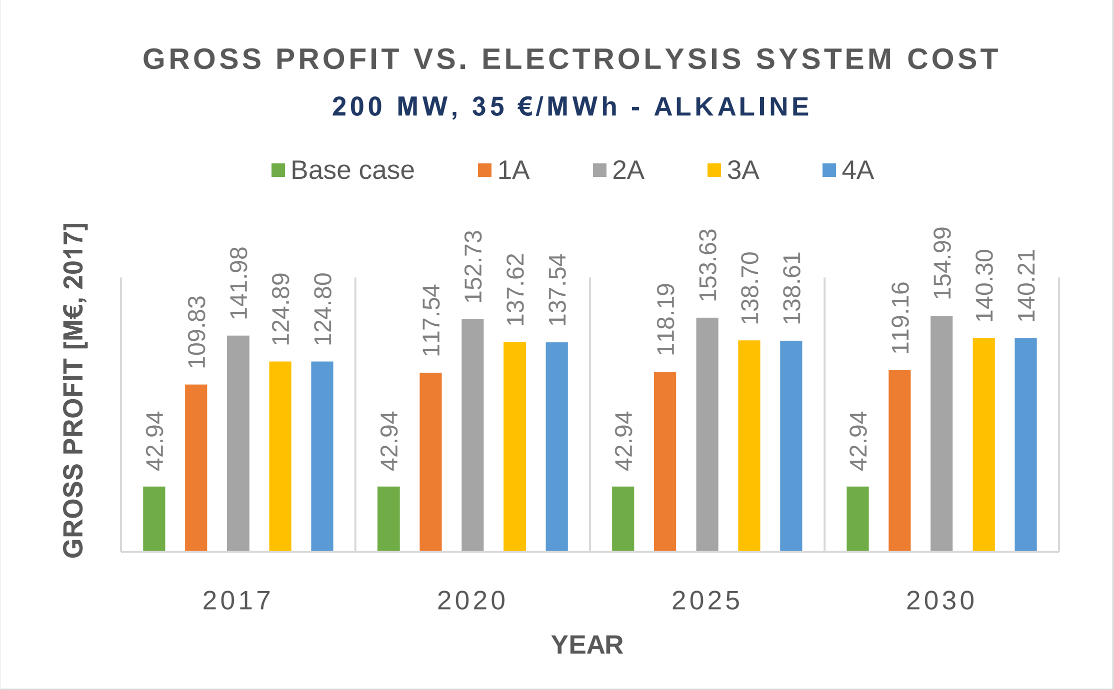

Figure 4.12 The relationship between gross profit and the year of a plant investment regarding the electrolysis system cost

5. CONCLUSION

The overall economic result shows that the process integration between conventional gasification and power-to-gas technology is profitable when using a specific operating parameters and economic variables as defined in this study.

According to the overall cost evaluation analysis, Configuration 2A is preferable for a wide range of electricity prices and can compete the base case configuration when the electricity price is up to 55 €/MWh. In addition, the result shows the trend that process Configuration 2A can still gain the profit even when the electricity price is up to 70 €/MWh.

Nevertheless, Configuration 3A and 4A tends to be rather preferable than Configuration 2A, due to its relatively higher profit, if it is operated when the electricity price is below 15 €/MWh. The reason is that a higher amount of H_2 is required in a Sabatier process for Configuration 3A and 4A, compared to Configuration 2 in order to produce a relatively similar amount of biomethane, leading to a higher revenue of by-product O_2.

To conclude, two main aspects to consider, when selecting the suitable process configuration for a process integration of biomass gasification and a power-to-gas technology, are an electricity price and an operational flexibility. First, Configuration 2A is preferable when the electricity in the market is relatively high (up to 55 €/MWh) and there is an abundant availability of H_2. Second, Configuration 3A is more suitable when the electricity price in the market is very low (below 15 €/MWh) which can be expected as a promising electricity price in the future market. In comparison with Configuration 4A, Configuration 3A is more preferable, since the cost difference regarding H_2 separation unit is not noticeable compared to the total capital cost; the operational flexibility is an additional benefit for choosing Configuration 3A.

From a present perspective, Configuration 2 is more preferable in terms of a current electricity price and a possibility to minimise CO_2 emissions, as recently concerned. This can be translated to a further economic perspective, depending on the CO_2 tax policies of the future.

5.1 Outlook and Future Works

The future works based on the finding of this work should be as following:

The process optimization on the performance of each process configuration regarding the electricity fluctuation behaviour should be of interest in order to predict the electricity production and operate the process according to a desired electricity price available in the market at a specific period of time.

The plant operation flexibility of each process configuration should be investigated in order to compare with the economic performance. It is always good to select the process design providing the highest profit. However, the operating flexibility and process performance are also essential to be considered since the profit might depend on these factors.

A detailed environmental assessment (e.g., in the form of cradle-to-grave life cycle assessment) should also be carried out, considering also future CO_2 emissions tax policies which can influence the production cost.

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APPENDIX A

Financial Parameters using in Economic Evaluation

Table A.1 Financial parameters and process variables used in economic evaluation

Financial parameters	Values	Note
Annuity factor (r)	0.12	*
Interest rate (i), [%]	0.1	*
Plant lifetime (n), [year]	20	*
Operating and Maintenance (O & M), [% of TCI]	4	[22]
Interest during construction, [% of TPC]	5	[22]
Annual operating hours [h]	8000	*
Biomass, [€/MWh]	18.516	[22]
Electricity, [€/MWh]	35	[25]
Oxygen, [€/kg]	0.161	[22]
Steam, [€/kg]	0.031	[22]
Methane, [€/kg]	1.766	[10]
LHV, $_{\text{Wet biomass}}$, [MJ/kg]	18.716	[10]
LHV, $_{\text{Dried biomass}}$, [MJ/kg]	19.34	[22]
LHV, $_{\text{Methane}}$, [MJ/kg]	50.016	[26]
LHV, $_{\text{Hydrogen}}$, [MJ/kg]	119.960	[26]

* Author's estimate

APPENDIX B

Description of Acronyms

Table B.1 The acronyms used in this study with the descriptions

Acronyms	Description
C	The cost values are in the currency of year 2017
C_o	The cost of a reference installed equipment of size S_o in 2017 million euros
S_o	The reference equipment capacity
S	The equipment capacity as a specified size
f	The cost scaling exponent
LCOF	Levelised cost of fuel, [€/GJ]
F	Feedstock cost (biomass residues), [€]
E	Electricity cost, [€]
C_p	Capital charges, [€]
O	Operating and maintenance costs, [€]
R	Revenue from selling by-products (steam and oxygen), [€]
M	Methane thermal output, [GJ]
r	Annuity factor
i	Interest rate, [%]
n	Plant lifetime, [year]
NPV	Net present value, [€]
M_c	Revenue from selling product methane, [€]
IDC	The indirect cost
PC	The project contingency
TPC	Total Plant Cost, [M€]
TCI	Total Capital Investment, [M€]
1A, 2A, 3A, 4A	Configuration 1,2,3,4 (integrated with alkaline electrolysis)
1PEM, 2PEM, 3PEM, 4PEM	Configuration 1,2,3,4 (integrated with PEM electrolysis)

Note: Acronyms used in process modelling scheme were retrieved from previous study [10]

APPENDIX C

Reference Equipment Costs

Table C.1 The reference equipment costs used in total capital investment

Equipment	So	unit of capacity	Co (M€)	IDC	PC	f	Notes
Biomass preparation and drying	64,600	kg/h, wet biomass	12.34	32%	-	0.77	b
Gasification	64,080	kg/h, dry biomass	38.78	15%	30%	0.7	c
WGS reactor	1377	Feed to gasifier, MW_{LHV}	12.96	15%	30%	0.67	c
Methanation	210	Methane, MW_{LHV}	33.43	15%	30%	0.67	c
Guard beds	260	Syngas, MW	6.17	10%	10%	0.85	c
Syngas cleaning	200000	Nm^3/h (NTP) input sourgas	58.33	15%	30%	0.63	d, c
Air separation unit (stand-alone)	76600	kg/h, oxygen output	37.86	10%	10%	0.5	e, c
Heat exchanger network	as retrieved*	Heat transferred, MW	as retrieved*	15%	30%	0.8	a, c, f
Syngas compressor	10	Compressor work, MW_e	5.14	15%	30%	0.67	g, c
Civil works (buildings and structures)	300	Feedstock, $MW_{LHV, AR}$	13.17	10%	30%	0.85	c
Sabatier reactor	as retrieved*	kg/h, gas inlet	as retrieved*	**		0.6	a, f, h, i
CO_2 Capture unit	as retrieved*	kg/h, gas inlet	as retrieved*	10%	30%	0.6	a, f, h, i
H_2O Separator	as retrieved*	kg/h, H_2O separator inlet		**		0.6	a, f, h, i
Alkaline Electrolysis							
• *2017*	1	Electricity consumption, MWh	0.84	15%	10%	0.93	a, c, j
• *2020*	1	Electricity consumption, MWh	0.61	15%	10%	0.93	a, c, j
• *2025*	1	Electricity consumption, MWh	0.59	15%	10%	0.93	a, c, j
• *2030*	1	Electricity consumption, MWh	0.56	15%	10%	0.93	a, c, j
PEM Electrolysis							
• *2017*	1	Electricity consumption, MWh	1.45	15%	10%	0.93	a, c, j
• *2020*	1	Electricity consumption, MWh	0.97	15%	10%	0.93	a, c, j
• *2025*	1	Electricity consumption, MWh	0.84	15%	10%	0.93	a, c, j
• *2030*	1	Electricity consumption, MWh	0.74	15%	10%	0.93	a, c, j
Compressors	as retrieved*	Compressor power, MW	as retrieved*	**		0.6	a, h
Turbine	as retrieved*	Delivered power, MW	as retrieved*	**		0.6	a, h
H_2 Separator	as retrieved*	Gas inlet, kmol/h	as retrieved*	15%	30%	0.95	a, k

Table C.1 The reference equipment costs used in total capital investment (continue)

Note	Detail
a	Author's estimate
b	Jin H. [27]
c	Hannula I. [22]
d	Liu et al. [28]
e	Larson et al. [29]
f	Retrived from Aspen Plus process modelling
g	Kreutz et al. [30]
h	Peter et al. [21]
i	Towlor and Sinnott [31]
j	Bertuccioli L. et al [24]
k	Marcoberardino G. D. et al [32]
*	The value was retrieved from Aspen Plus process modelling and scaled-up regarding the specific production capacity by using the literature references and author’s estimate.
**	The typical factors for equipment capital cost estimation was used to account the indirect cost (including offsites, design and engineering, and contingency), suggested by Towlor and Sinnott [31].

APPENDIX D

Detail Process Explanations

D.1 Biomass Preparation and Drying

The biomass preparation and drying are an essential stage prior the gasification process since the biomass particle size is usually large, in the range of 10 - 80 mm [33], and it contains high initial moisture content, typically between 50 - 60% of wet basis [11]. The higher percent moisture contains in biomass, the higher heat demand for gasification is required. As a reason, the biomass is normally dried to achieve a moisture content below 15% on wet basis [11]. The typical values of dried biomass fuels are in the range of 15 - 20 MJ/kg dry ash free [12]. Since the moisture content of biomass can be reduced to 10 - 15% wet basis when a low-temperature drying (at below a water boiling point) is used [13], the drying technique that is applicable for low-temperature drying is preferred. In addition, the capital cost of gasifier and other ancillary equipment cost can be reduced by using biomass drying since the biomass chip size can be reduced during the drying process, leading to a decrease in equipment dimensions [11]. Moreover, the combustion control issues can be minimised by using fuel with low moisture content [13]. On the other hand, an intensive energy consumption is required in biomass drying, consequently, it emphasizes that the use of waste heat at low temperature can be beneficial to the overall process efficiency.

The most common biomass dryers are rotary dryers, fluidized bed dryers, and belt dryers. According to the advantage of using low-temperature heat in biomass drying, the belt dryer is comparatively preferable since it is suitable to operate at low temperature in range of 60 - 200 ºC. However, the preferable range of heating temperature is between 120 - 150 ºC in order to prevent the risk of explosion at elevated temperature and biofuel devolatilization at very low temperature [11]. Biomass is transferred to the sieve conveyor and transported along the dryer. The drying medium (i.e. hot air and steam) is blown through the conveyor and the biomass bed by fans. Conventionally, the height of biomass bed is in the range of 2 - 30 cm depending on the type of biomass [11]. As the low-temperature heat can be used in the belt dryer, the risk of explosion and the emission of pollutants for such a VOCs, due to a high temperature heating, can be minimised [34]. However, the higher retention time and larger installation area are required for belt dryer due to its low-temperature heating [11].

D.2 Gasification

Gasification is a thermochemical process converting any solid fuel (i.e. biomass) into gas products, in the presence of a gasifying agent such as air, oxygen, or steam. The produced gas mixture consists of CO, H_2, CO_2, H_2O, CH_4, and small quantities of hydrocarbons, tars, impurities, as well as N2 when air is used as a gasifying agent in the process [35]. Gasification takes place at temperatures between 500 - 1400 ºC, and at atmospheric or up to 33 bars [5]. Gasification occurs in endothermic reactions via two typical processes: the auto-thermal process (direct gasification) and the allothermal process (indirect gasification), classified by how heat is supplied to the process [36]. In the former type, heat is generated inside the reactor by partial combustion of the feedstock, while the heat is produced in a separate reactor in the latter, meaning that mass and heat is transferred between the reactors [36]. The conceptual idea of direct and indirect gasification processes is illustrated in Figure 1.1.

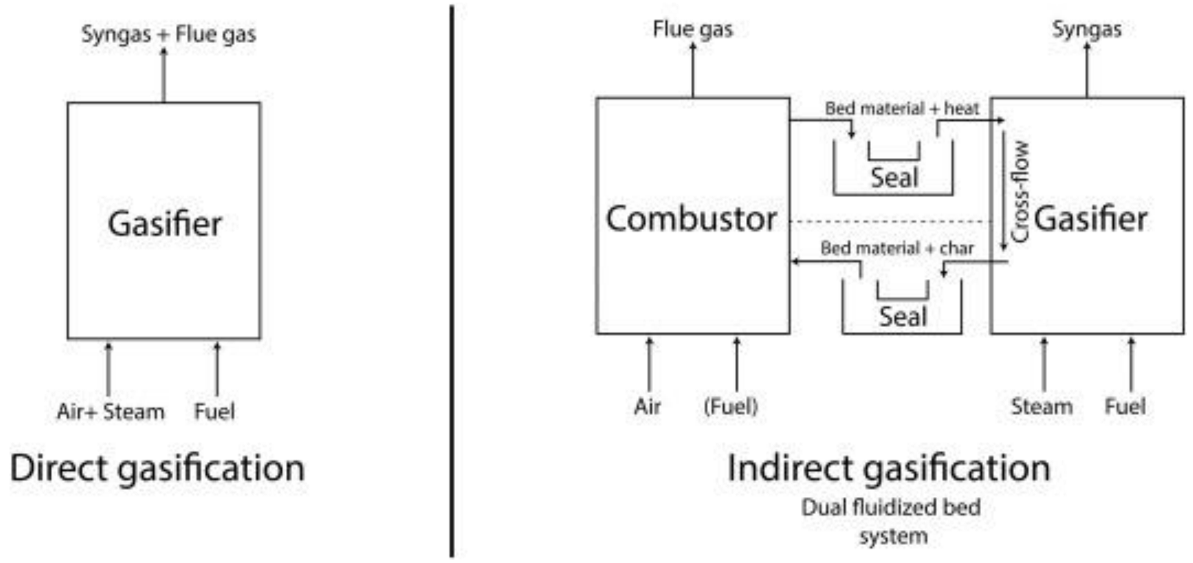

Figure D.1 Direct and indirect gasification processes [8]

In addition, there are three basic types of gasifiers: fixed bed, fluidized bed, and entrained flow gasifiers [9]. The fluidized bed gasifier is a promising technology to be used in industrial-scale biomass gasification and seems to be the most suitable design for biomass gasification since it can overcome the drawbacks of fixed bed gasifier in terms of mixing, retention rate, and the limitation in production capacity [9]. However, there is only one biomass gasification located in Gothenburg which is operated in a commercial scale, recently [3]. In case of entrained flow gasification, although it provides higher performance comparing to fluidized bed gasification, it is considered as less attractive to be used in biomass gasification. For example, the common difficulties in operating entrained flow gasifiers are an economical aspect of biomass particle size reduction, the corrosion of the reactor lining caused by molten slag from biomass, and inherent equipment size limitations [9].

D.3 Syngas Cleaning

The syngas cleaning (or syngas conditioning) is required to eliminate some impurities containing in the gas mixture that could affect the downstream process or the product gas utilisation. The gas treatment also helps remove undesired gas compounds and adjust the gas component ratio to a proper value according to the downstream process requirements [15].

The composition of undesired gas compounds and other impurities depends on the type of feedstock and gasification process type. Typically, the undesired gas compounds and impurities include: particulate matters, sulphur compounds, nitrogen compounds, halogens, volatile metals, tars, hydrocarbons, and carbon dioxide [17].

D.3.1 Particulates

The particulate matters containing in the gas mixture typically come from the ash and dust of feedstock, unconverted carbon, soot, and bed materials when the fluidised bad gasifier is used. The conventional particulates removal technologies are cyclones, barrier filters, electrostatic filters, and scrubbers. Cyclones, moving bed filters, and ceramic filter candles are suitable for a high temperature gas separation, while scrubbers and wet electrostatic filters are suitable to a low temperature separation [15].

Cyclones are appropriate for a first stage particulates separation and generally used for a wide range of temperature. However, the separation efficiencies are low, and a high amount of particle content normally remains in the gas mixture after separation [17]. Therefore, the particulates removal technology used in this study was a ceramic filter since the gas outlet of the gasification was at high temperature and the separation efficiency of ceramic filter is relatively high comparing to cyclones.

D.3.2 Tars

Tar is an organic compound contained in the product gas mixture after gasification. Tar can cause the risk and difficulty in the downstream process since it may condense on the heat exchanger surface, poison catalyst, leading to a high process maintenance cost. The conventional tar removal technologies are a high temperature destruction, a catalytic destruction, and a physical removal with organic washing liquid [17].

The physical techniques of tar removal (i.e. liquid scrubbing) are appropriate for a low-temperature gas inlet where the product gas from gasification has to be cooled down prior the scrubber, resulted in a loss of overall thermal efficiency and a need in an additional treatment for a waste liquid stream. Therefore, a catalytic reforming is considered as a better alternative to decrease tar concentration in the gas product mixture since tars can be destructed at high temperature and converted to light gases without generating a waste liquid stream [37].

D.3.3 Steam and Moisture

After tars removal, the syngas is normally proceeded to a water scrubber to cool down the gas mixture followed by condensing the moisture to dry the gas mixture at the specific requirement of the downstream process [22].

D.3.4 Acidic Gases Removal

During gasification, sulfur containing in biomass feedstock can be converted to hydrogen sulphide or sulfur oxides which further contain in the gas product mixture. Although sulfur in biomass is relatively low, normally in between 0.3 - 0.4% or less than 0.1% of sulfur by weight, comparing to coal, it can cause problems in the downstream process, even at a low concentration, especially when the catalysts are used in the synthetic gas system. As a reason, the acidic gases removal is required when the process is dealing with catalyst activities [19].

The acid gases (H_2S and CO_2) removal techniques are normally divided into wet and dry processes. The wet processes (or absorption) are typically MEA (Monoethanolamine) and DEA (Diethanolamine) processes, MDEA (Methyldiethanolamine) processes, and rectisol process. The dry processes (or adsorption) are mainly adsorption with metal oxides, zeolites, and activated carbons [17].

In this study, rectisol process was used as an acidic gases removal for upstream process since it has high selectivity for H_2S versus CO_2, therefore a rich H_2S off-gas stream can be obtained even if there is a low concentration of H_2S containing in the gas mixture inlet. Moreover, the desired gas loss in the off-gas stream can be minimised since the solubility of H_2, CO, and CH_4 in the liquid absorbent are low [17]. For the CO_2 removal in the downstream process, an absorption process using MEA solvent was used. A detailed explanation of CO_2 removal in downstream process is presented in section 2.6.2.

D.4 Water-Gas-Shift Reaction

The composition ratio of H_2 and CO in the gas mixture has to be adjusted regarding to type of biomass feedstock in order to obtain the desired gas ratio of H_2/CO prior to a methanation in the downstream process. The ratio of H_2 and CO can be adjusted by water-gas-shift reaction, as shown in Equation 2.1 [17].

$$CO + H_2O \rightarrow CO_2 + H_2 \qquad (2.1)$$

The water-gas-shift equilibrium depends on temperature while it is independent on industrial pressure range. To forward an equilibrium reaction to favour the production of hydrogen, the reaction should be operated at low temperature with a catalyst. However, the selected operating temperature depends on the required ratio of H_2 and CO. When more CO composition in the gas mixture is needed, the reaction has to be operated in a high-temperature shift of 300 - 510 ºC, with copper-promoted catalysts. On the other hand, a low-temperature shift of 180 - 270 ºC, with copper-zinc-aluminium-oxide-based catalyst, is prefer when more hydrogen production is required [17].

D.5 Methanation Process

The produced gas mixture after gasification consists mainly of carbon monoxide, hydrogen, methane [16]. Since the gas mixture contains an uneconomically low quantity of methane, the amount of methane is normally increased by using the technique called methanation process. Methanation is a primary technique used to produce methane from the reaction between carbon monoxide and hydrogen gas containing in the produced gas mixture of gasification, as shown in Equation 2.2 [18].

$$CO + 3H_2 \rightarrow CH_4 + H_2O \qquad (2.2)$$

The synthetic gas methanation processes are classified into two types: fixed-bed and fluidized-bed methanations. The fixed-bed methanation is mostly used since it can overcome the drawbacks of using fluidized-bed reactor in terms of a catalyst abrasion and a scale-up difficulty. As a methanation process is a drastically exothermic reaction, several adiabatic fixed-bed reactors are generally used to control an increase in temperature when the reaction occurs. The gas mixture is cooled down by the intermediate cooling between each reactor to limit the maximum temperatures [38]. Apart from the methanation of CO and H_2, the methanation of CO_2 presents in the gas mixture can also occur regarding Sabatier reaction (see section 2.6.1). Nevertheless, the methanation of CO is more favoured in high temperature since the heat released during the reaction influences the reverse water-gas-shift reaction, leading to an increase in CO concentration in the gas mixture [39].

After the methanation process, the gas mixture contains mostly methane and carbon dioxide as well as low amounts of steam. To obtain a market purity standard of biomethane, carbon dioxide in the gas mixture outlet has to be removed for the conventional production plants.

D.6 Sabatier Process

The gas mixture contains mostly biomethane and carbon dioxide (produced mainly in water-gas-shift reaction) after the methanation process. To purify the product biomethane, carbon dioxide in the gas mixture can be handled in two ways: to be separated and release to the atmosphere or to be injected with H_2 to Sabatier process in order to enhance the yield of product methane. By handling CO_2 in the latter alternative, not only the yield of biomethane can be increased, but also the need for carbon dioxide separation and the amount of CO_2 emission can be decreased. However, some additional amount of pure hydrogen gas is required in the process, which is considered as an operating cost penalty.

D.6.1 Sabatier Reaction

The Sabatier process is an exothermic reaction where CH_4 and H_2O are formed by the reaction between CO_2 and H_2 in a presence of catalyst, as shown in Equation 2.3 [8].

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O \qquad (2.3)$$

To favour a higher conversion of CO_2, the Sabatier process is recommended to operate at low temperature, approximately between 250 - 400 ºC. The catalysts that are mostly used in the Sabatier process are Ni and Rh which have been proved in operating at industrial scale [40].

The Sabatier reactors recently used are a conventional packed-bed reactor and a microchannel reactor which provides better mass and heat transfer between reactants and channel wall, leading to a higher CO_2 conversion [41]. Due to its substantial exothermic reaction, the reaction temperature has to be controlled by designing at least two Sabatier reactors with an intermediate cooling, similarly to the methanation process as described in section 2.5. In this study, the Sabatier reactor invented by Turbo SE and MAN Diesel [19] was used in the process. It was designed as a single reactor that has two regions with a shell-and-tube reactor filled with catalyst pellets. The cooling agent flows on the outer surface of the tubes to control the reaction temperature [38].

D.6.2 Carbon Dioxide Capture

The conventional carbon dioxide separation technologies recently used are chemical absorption, physical absorption, cryogenic methods, membrane separation, and biological fixation. The most effective and often-used method is a chemical absorption using aqueous monoethanolamine (MEA) as an absorbent due to its high reactivity with CO_2, low solvent cost, and its regenerable property [42].

The CO_2 absorption process using amine solvent includes two unit operations: absorber (CO_2 absorption) and stripper (CO_2 desorption), as illustrated in Figure 1.2 [43]. Carbon dioxide in the gas mixture is feed into the absorber column and captured by aqueous MEA absorbent. The clean gas leaves at the top of absorption column while the rich amine solvent dissolving CO_2 is sent to the stripper from the bottom stream of the absorber. After a rich amine stream is cooled down prior to the stripper, CO_2 is then desorbed out of the liquid solvent and leaves the at the top of stripping column while the amine solvent is regenerated and recycled back to the absorber.

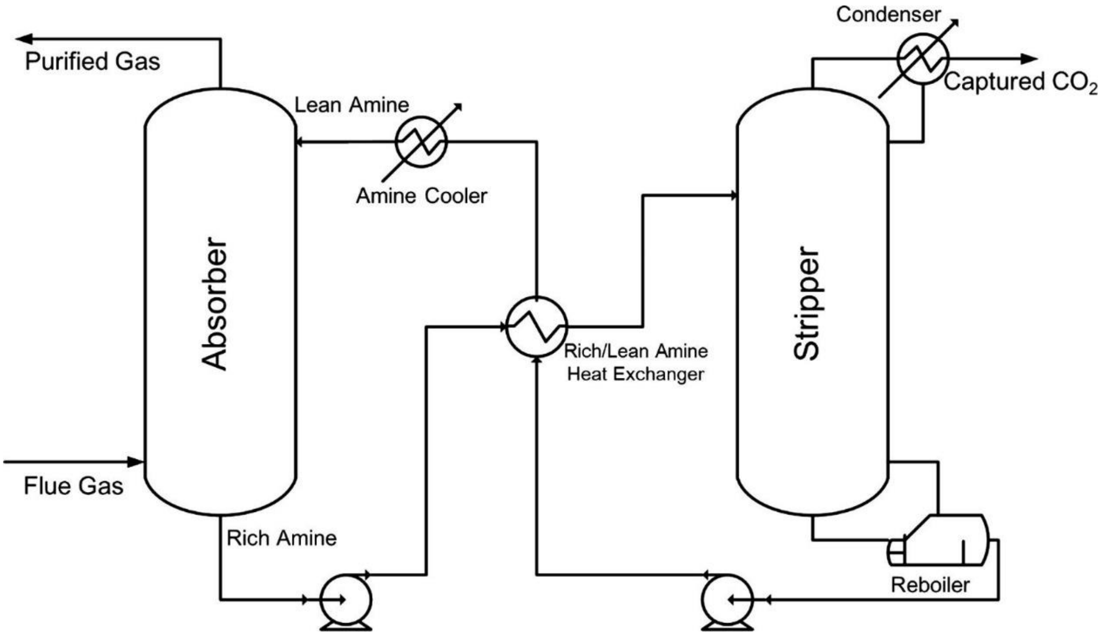

Figure D.2 Conventional flowsheet for amine-based CO_2 capture [43]

D.6.3 Hydrogen Separation

The hydrogen containing in the gas mixture after Sabatier process has to be removed in order to meet the Wobbe Index standard of biomethane production. A well-established process for hydrogen separation is a pressure swing adsorption (PSA) which is widely used in industrial level. PSA technology provides benefits in terms of low energy consumption, reliability, cost effective, and high purity level of hydrogen. The general principle of PSA technology is that hydrogen is a very high volatile compound with low polarity, therefore it is hardly adsorbed by adsorbent materials comparing to other molecules for such CO_2, CO, and hydrocarbons. These particular properties of hydrogen make PSA system highly efficient technology to use for hydrogen separation [44].

D.7 Power-to-Gas Technology

Power-to-gas concept is a technology that converts electricity power into a valuable gas that could be stored and used as a fuel. In gasification process, a power-to-gas concept is applied to produce H_2 and O_2 by water electrolysis, as shown in Equation (2.4), where H_2 gas is then proceeded to react with CO_2 in a Sabatier reactor to produce biomethane [45].

$$2H_2O \rightarrow 2H_2 + O_2 \qquad (2.4)$$

The electricity power that used to convert water into H_2 and O_2 could come from the intermittent renewable energy sources for such wind and solar energy, but more researches are required to ensure the utilisation of these energy while maintaining the stability of the grid. By apply the power-to-gas technology application to gasification process, it can not only make use of existing gas network [20], but it can also benefit the plant when the produced O_2 gas from electrolysis can be used as a gasifying agent for a direct gasification.

D.7.1 Electrolysis System

The electrolysis system is the main process equipment for power-to-gas technology. The water electrolysis technologies that are mature and recently available in the markets are alkaline electrolysers (liquid electrolyte) and polymer electrolyte membrane (PEM) electrolysers [45].

The advantages of alkaline electrolysers are that they are cheap, have long life spans, and that they are availability for large plants. However, there are drawbacks regarding its operation phase such as operating pressure limitation, high maintenance cost, and high environmental risk. The PEM electrolyser is developed to withstand high operating pressure, have cold-start time and to be operated in wide-range of production capacity, consequently, higher investment cost is needed [46]. To compare the capital investment for the electrolysis system, the predicted capital cost for alkaline and PEM systems was shown in Figure 1.3 [24]. From the figure, the capital cost of PEM electrolysis is predicted to reduce gradually by 2030 while the capital cost of alkaline electrolysis tends to decrease up to 2020 and stays constant at a specific price. This information was used as a base of capital cost estimation in the following sections.

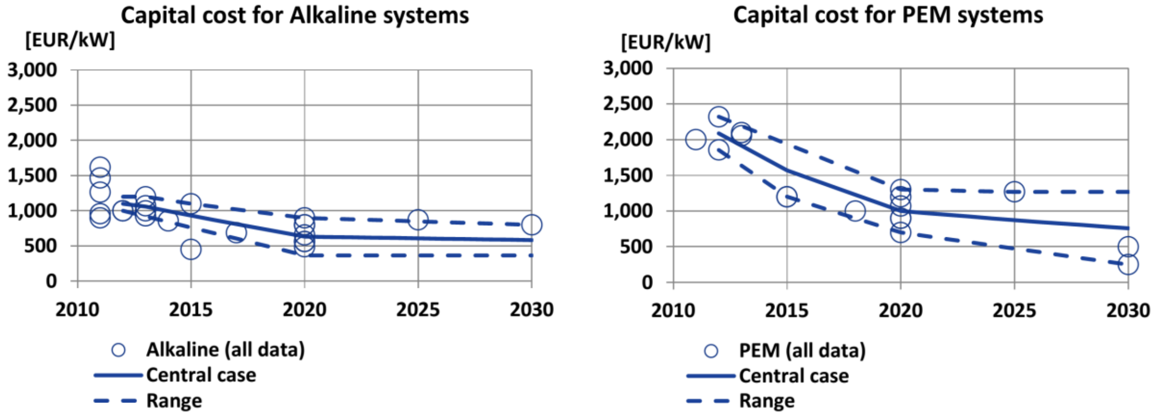

Figure D.3 Capital cost reduction trend lines for alkaline and PEM electrolysers [24]