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Techno-economic assessment of biomethane production from syngas derived from biomass gasifier

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Summary:

Biomethane is one of the renewable fuels widely utilized for various purposes, including transportation, heating, and power production. Growing interest is currently seen in this biofuel production utilizing waste material such as agricultural residual, sludge and municipal solid waste in anaerobic digestion. However, there is a limitation in using entire organic biomass in anaerobic digestion due to lignocellulose mass and fibrous material, which is hard to digest by microbes. Thus, gasification could be a supplementary technology to utilize residual, digested, and contaminated waste. The coupling of gasification and anaerobic digestion has been tested in lab scale reactor to valorize the waste as biofuel. However, techno-economic evaluation is essential to upscale the technology. Thus, this master's thesis aimed to perform the techno-economic of coupling gasifier and anaerobic digestion to utilize syngas for biomethane production. Three potential scenarios were evaluated for their techno-economic viability. Scenario 1 is based on biomethane generation through syngas fermentation coupled with gasification. Combination of hydrothermal gasification, syngas fermentation, and a steam addition as a hydrogen source make up Scenario 2. In scenario 3, an electrolytic unit was added instead of steam (in scenario 2) as a hydrogen source to produce biomethane. The scenarios' efficiencies range from 13% to 35%. The maximum energy efficiency, 35%, was found in Scenario 3. Furthermore, scenario 1 (237 NOK per litre) was followed by scenario 2 (164 NOK per litre) and scenario 3 (120 NOK per litre) as the scenarios in which the minimum selling price of biomethane decreased. Due to its undiscounted net present value of 76 million NOK compared to the other scenarios, scenario 3 is the most profitable based on the discounted cash flow analysis results. According to sensitivity analysis, the cost of labour and utility had the most impact on the minimum selling price of biomethane.

Preface

This master thesis was written in the spring of 2023 as the final project to attain MSc. in Energy and Environment Technology at the University of South-Eastern Norway (USN).

A techno-economic evaluation of two stage (gasification-AD) biomethane production process has been performed. In addition, economic methane production rate scenario for up scaling the technology has been proposed.

The software utilized for calculations and simulations in this thesis is Excel and Aspen Plus.

I want to thank my supervisor Nabin Aryal and my co-supervisor, Lars Erik Øi. I am so grateful for all their help and input during this project. Having supervisors with such knowledge and interest in the subject was motivating.

Last but not least, I want to sincerely thank my wife for her encouragement and support this semester. Without her continuous assistance and support, I could not put out the necessary efforts. I'm grateful.

Lysaker, 12.05.2023

Milad Sahraro

Contents

Nomenclature

AD	Anaerobic digestion
BFB	Bubbling fluidized bed
BM	Bed material
BT	Bed temperature
BTA	Batch Reactor autoclave
BTR	Batch type Reactor
CAPEX	Capital expenditure
CB	Catalyst to biomass ratios
CCE	Carbon conversion efficiency
CEPCI	Chemical engineering plant cost index
CFB	Continuous fluidized bed
CSTR	Continuous stirred tank reactor
DBEP	Discounted break even period
DC	Direct cost
DCFA	Discounted flow analysis
DCFRR	Discounted cash flow rate of return
EFB	Empty fruit bunch
EPC	Equipment purchase cost
ER	Equivalence ratio
EXP	Experimental
F	Feedstock
FCI	Fixed capital investment
FFA	For first agent
FIB	Fixed bed
FLB	Fluidized bed
FOC	Fixed operating cost
FR	Feed rate
FS	Feedstock size
FSA	For Second Agent
GA	Gasifying agent
GH	Gasifier height

GT	Gasifier types
HRT	Hydraulic retention time
ID	Gasifier internal diameter
IGAD	Integrated gasification and anaerobic digestion
INC	Indirect cost
IRR	Internal rate of return
IT	In total
LCA	Life cycle assessment
LCFA	Long chain fatty acids
MPa	Mega Pascal
MSP	Minimum selling price
MSW	Municipal solid waste
NA	Data were not available
NOK	Norwegian krone
NPV	Net price value
NRR	Net rate of return
OC	Optimum Conditions
OLR	Organic loading rate
OPEX	Operating expenditure
OSW	Organic solid wastes
PAH	poly aromatic hydrocarbons
PBP	Payback period
PEM	Polymer electrolyte membrane
RHS	Right hand side
RN	Reference Number
SB	Steam to biomass ratio
SIM	Simulation
TAN	Total ammonia nitrogen
TEA	Techno-economic analysis
TG	Total gas
USD	United states Dollar
VFA	Volatile fatty acids
VOC	Variable operating cost

WBM	With bed material
WC	Working capital
WOB	Without bed material

1 Introduction

The most common approach to use for determining the viability of a new idea or project is a techno-economic analysis (TEA). Typically, this calls for a process to be carried out at a large-scale facility, particularly when no prior knowledge about comparable industrial facilities is available.

Due to potentially renewable antecedents and environmentally friendly manufacturing methods, biofuels are a practical substitute for fossil fuels [1], [2]. Biomethane is a widely used biofuel[3]. It can be made through the fermentation of syngas with bacteria or through the fermentation of sugar with yeast [4]. Biodiesel, another type of biofuel used in the transportation industry, is made from a variety of feedstock, including both edible and inedible oils, algae, and lignocellulosic biomass [5]. It is created by transesterifying short-chain alcohols, catalysts, and vegetable and animal fats. Because transesterification produces a substantial amount of crude glycerol, it is undesirable from an economic and environmental aspect [6]. Biomethane, one of the principal gaseous biofuels, can replace natural gas in power and heat production. It can be created by the anaerobic breakdown of organic material whenever methanogenic bacteria are present [7]. The Sabatier reaction (Eq. 1), which can occur by chemical or biological synthesis, can also yield biomethane [8]. Methane is produced by biological metabolism, which recycles CO_2 through syngas fermentation [9], [10].

$$CO_2 + 4H_2 \xrightarrow{yields} CH_4 + 2H_2O \tag{1}$$

The current study suggests a novel conceptual framework that combines biomethanation, syngas fermentation, and hydrothermal gasification. The economic viability of the sequential biomass-to-gas and gas-to-liquid process was further assessed using a thorough techno-economic and sensitivity analysis.

The primary purpose of this master thesis is

1. To perform the techno-economic evaluation of a two-stage (gasification-AD) for biomethane production.

2. To propose the economic methane production rate scenario for up-scaling the technology

The main tasks are:

i) Designing of a process framework for biomethane production

ii) Establishing the mass and energy balances based on the designed framework,

iii) Total cost estimation based on the capital expenditure (CAPEX) and operating expense (OPEX) and

iv) Performing sensitivity analysis.

The complete task description is in Appendix A.

Chapter 2 includes a literature review of effective parameters on gasification and anaerobic digestion processes.

Chapter 3 provides an overview and methodologies which are dominant on simulation of different scenarios in Aspen Plus and TEA analysis.

Chapter 4 reports the results of simulation an economic evaluation of those scenarios.

Chapter 5 contains a discussion about the results in previous chapter.

Chapter 6 includes conclusion and suggestions.

2 Literature review

2.1 Bio fuels

Recently, biomass consumption as a sustainable energy source has attracted considerable attention worldwide. According to McKendry [11], biomass consumption has been predicted to provide about 10-14% of the world's power demand. One of the significant advantages of biomass over fossil fuels such as oil, gas and coal is its steady feedstock supply [12]. In addition, fossil fuels have ambiguity in price and supply levels. Moreover, the broad use of fossil fuels has consequences such as green-house effect and results in climate change [13]. Life Cycle Assessment (LCA) analysis shows that biomass is not only a CO_2 neutral resource [14] but also has zero CO_2 net emission energy [15]. Compared with biomass, fossil fuel consumption causes extra irreversibility. It extracts the carbon locked underground and injects it as CO_2 into the atmosphere. Meanwhile, forests (biomass sources) are sustainably managed [16].

2.1.1 Advantages of Biofuels

Considering the environmental effects, sulfur concentration is insignificant in biomass fuels which is responsible for generating ash. Sulfur is consequently generating considerably less air emissions in comparison with fossil fuels. So, burning biomass doesn't play role at sulfur-dioxide emissions, which is cause of acid rain and the amount of ash which is produced is usable in farms as a soil additive [12]. Another noticeable point is that biomass consumption will reduce the amount of waste on the landfills. This waste has a considerable impact in disposal problems, especially in urban areas. The economic advantage of biomass usage as an energy source is that independence from fossil fuels increases, and economic pressures of importing petroleum products will significantly reduce [17].

2.1.2 Biomass conversion technologies

There are four methods for biomass conversion technologies. Direct combustion processes, biochemical processes, agrochemical processes and thermochemical processes [12].

The most dominant method for generating energy from biomass is direct combustion, in which the produced heat can be utilized for different proposes such as heating and generating electricity. Biochemical processes consist of AD and alcoholic fermentation. At the same time, an example of agrochemical process is the extraction operation which is fulfilled mechanically. Thermochemical process can be used for all sorts of biomasses, especially for low moisture, herbaceous and woody materials [17].

2.1.3 Biomass types

In general, biomass can be divided into five main categories. The first one is virgin forest wood which originates from the waste of forest products such as sawdust, wood chips, wood pallets, and palm kernel shells. The second group are energy crops, which yield high amounts of crops

grown specifically for energy applications. For example, Hybrid eucalyptus, and Napier grass, which are not suitable for agriculture, are categorized in this group.

The third group is agricultural residuals, such as bagasse from sugarcane and straw [18]. The fourth group is food residuals which consist of food and drink manufacturing residuals, animal fat, etc. The final group is residuals from industrial or municipal waste and manufacturing processes [19]. The first group is the main source which contributes (64%) of biomass and MSW is (24%) agricultural waste (5%) and landfill gases (5%) [17].

2.2 Biomass gasification process

Syngas is one of the important transitional gases in the chemical processes such as Fischer-Tropsch liquids, ammonia and methanol production [20]. It can be utilized as a source of pure hydrogen and CO [21]. Recently, hydrogen is the favorable energy source consumed in fuel cells and combustion engines [22]. Generally, syngas is produced in thermochemical processes from biomasses. These processes can be pyrolysis, gasification, and combustion or reforming processes. However, syngas can be produced in biological processes such as direct and indirect biophotolysis, biological water-gas shift reactions, photo-fermentation and dark-fermentation [13].

Among thermochemical processes, gasification is the most efficient process to produce hydrogen out of biomasses. Although, the firs gasification plant was constructed in 2001 but, syngas and bio-mass gasification were both known since WW II [23]. Numerous kind of carbon-based feedstock can be used in gasification process. For example, industrial waste, biomass, petcock, petroleum, coal and natural gas are some of them [24]. As shown in Figure 2.1, coal, petroleum, natural gas, and petcocks have 51%, 25%, 22% and 1%, respectively. Number of plants which uses those feedstock are 53 and 56 plants for coal and petroleum, respectively [24].
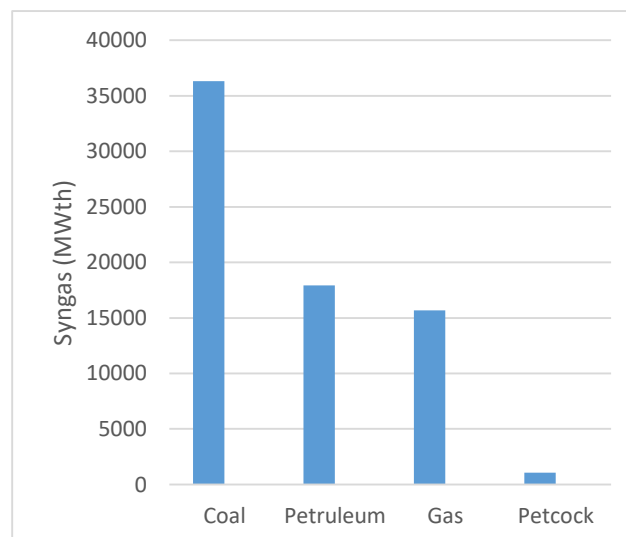
Fig 2.1. Gasification operating capacity by feedstock in the world [12]

Biomass gasification can be described as a process in which carbonaceous material converts into a gaseous product or synthesis gas that dominantly includes H_2, CO, CO_2, H_2O, CH_4, C_2+ (higher hydro-carbons), and N_2 [24].

Gasifying agents such as air, pure oxygen, steam, or mixture of these gases, is one of the necessities in the gasification process, which is conducted at high temperatures between 500 and 1400°c where the pressure range is between 1 to 33 bar [25]. Peres et al [13] state that gasification is suitable for biomasses with less than 35% moisture content. Otherwise, great amount of power loss will be expected in the process where as the moisture content in the feedstock is between 25% to 60%. Nooruddin [26] recommend to dry and preheat the bio-mass to reach the moisture content between 10% and 20% before being introduced to gasifier.

First stage for gasifying the biomass is devolatilization. During this process char and volatiles will be produced and then char will be gasified and volatiles and then it will be cracked and reformed. Adding steam in this step will increase the reforming reactions for volatiles and char gasification. As a result, light gases such as H_2, CO and CO_2 will be generated [27]. In this stage, produced gases can be utilized either for heating and electricity generation or for being further synthesized and stored as a fuel. On the other hand, this gas can be utilized as a raw material for high-value chemical productions like olefins and formaldehyde [28]. Gasification steps has been shown in Figure 2.2.
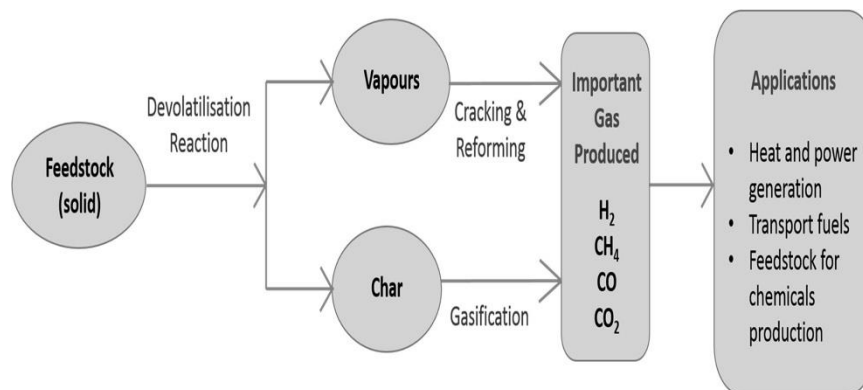
Fig 2.2. Process flow for gasification and its application [12]

There are some parameters which significantly affect the rate and composition of the produced gases. For example, type of gasification reactor, operating conditions, properties of input masses and output conditioning are the main ones. Where to inject the feedstock and fuel is a key factor in defining the type of gasification reactor. Pressure, temperature and residence time are related to operating conditions which control the reaction of gasification [29]. Gomez et al. [30] has investigated effect of utilization of different types of gasifiers, called entrained flow, fixed bed and fluidized bed.

Paula et al. [13] have studied the agents' effect in gasification like: air, oxygen, steam, CO_2, or mixture of these agents. They have also investigated operating conditions that consist of the temperature, pressure, catalyst and the equivalent which affect the amount of yields and composition of products.

Input masses are related to the raw materials' properties. (type and components) [31]. Output conditions of gas include cooling and disposal of specific matter and tar from produced gas [29].

2.3 Gasification performance

Within the gasification process, syngas generation is influenced by several parameters. In order to study and design the most efficient gasification system, factors such as types of gasifier and processing conditions (for example: temperature, gasification agents, size of biomass particles, and bed material type), should be investigated thoroughly.

2.3.1 Gasifier types effect

Gasifiers are categorized in various types. Fixed bed, moving bed and fluidized bed are the main groups of gasifiers. Gasifiers are divided into different types based on the flow direction and orientation of biomass and oxidant flow into the reactor and also the way that heat is transferred to vessel. The first group is fixed bed gasifier, this group can be divided into updraft and down drift gasifiers. In this category, biomass bed position is fixed and the combusted gases pass through it. In another case, bed can move along the reaction, or can move mechanically. The fixed bed updraft is also called counter current gasifier. This type of gasifier is the simplest and the oldest type, and also has cheapest concept among other gasifiers. In this type biomass is entered at head part of reactor and the gasification agent is introduced from the bottom. On the other hand, both the oxidant and the bio mass are entered from the head of reactor, but all the mechanical configurations are as same as the updraft gasifier [24]. The downdraft gasifier is also called co-current gasifier.

The second group (fluidized bed gasifiers) is divided into two common categories that are bubbling and circulating fluidized- bed gasifiers. In bubbling fluidized bed, commonly, there are fine and inert particles of sand and Aluminum.

In this type gasification agent is pumped in a way to pass from the inert particles. The speed of agent increased in a way to pass from the inert particles floated against its weight. At this gas velocity (minimum fluidization), channeling and bubbling of the gas happens. In this situation, particles remain in the reactor and appear to be in a "boiling state". On the other hand, in circulating one, operating gas velocity is higher than the minimum velocity. In this case, particles will be mixed with the gas stream and flow to top of gasifier and finally become separated by a cyclone, and then it enters the same cycle again [25].

Usually, the particles which are utilized in circulating type, are smaller than bubbling one [31]. Circulating one also had higher conversion rate and efficiency in comparison with bubbling one [32]. Among current commercial or near-commercial gasification technologies, directly heated bubbling fluidized bed (BFB) gasifiers are the most widely used one which operates in a broad span of temperatures, pressures and throughput [24].

Some investigations have studied advantages and disadvantage of fixed bed gasifiers in comparison with fluidized bed ones. Warnecke et al. [33], has carried out this comparison according to some aspects such as economic and environmental impact, use of material, energy consumption and the technology. They have stated that there were no significant advantage of each one compared to the other. Alauddin et al. [34]have shown that there are some drawbacks

for fixed bed gasifiers. They claimed that. There is a huge amount of tar and char production because there is a low and non-uniform heat and mass transfer between the gasifying medium and solid biomass inside the gasifier.

On the other hand, (FLB) gasifiers have high mixing and gas-solid contact, which improves the reaction and a conversion rate. In addition, a lower portion of tar in the product gas will be obtained when bed material play role as heat transfer medium and catalyst. As a result, in such a conditions, the quality of gas will be improved. These improved heat and mass transfer properties in FLB gasifiers will increase the throughput of gasifier for usage of various wastes with different compositions and thermal properties [34], [36].

2.3.2 Temperature effect

Temperature is one of the critical parameters in the gasification process. In addition, the temperature often regulates the thermal degradation process inside the reactor, thereby changing syngas composition. The lower temperature often supports to build of the tar and char fraction; thus, CO_2 composition is often high compared to CO, CH_4 and H_2. However, higher (around 800-1000) temperate thermal reactions stimulate tar cracking, and thermal reactions often provide more CO, CH_4 and H_2 gases compared to the CO_2 counterpart [35]. Therefore, a higher temperate greater than 1300°C could lead to low CH_4 composition. To support such argument, one of the studies from Mareno et al. [25] stated that when the operating temperature is greater than 1300°C, resulting the formation of higher hydrocarbons or tar, little to no methane, and a maximized generation of CO and H_2, was occurred. A similar observation was reported that there is an increase in total gas yield, and H_2 and CO composition during the temperature enhancement from 600 to 800°c,. However, they indicated that char and tar decreased during this temperature increase [37]. Mohammed et al. [15] also gasified the fluidized bed by empty fruit bunch (EFB) as biomass raw material and air as an agent in a temperature range between 700 and 1000°C. They indicated that during the enhancement of temperature from 700 to 1000°C concentration of CH_4 and H_2 increased from 5.84 to 14.72 and from 10.27 to 38.02 vol.%, respectively. They stated that during that process concentration of CO enhanced while the CO_2 concentration reduced.

In fluidized bed reactor, Lv et al. [38] investigated temperatures between 700 and 900°C. They utilized sawdust as raw material and showed that by increasing temperature in this range, higher amount of H_2 will be obtained and in contrary less CO_2, CO and CH_4 will be produced. They stated that the carbon conversion efficiency (CCE) also increased from 700 to 750°C. On the other hand, they indicated that when the temperature increases from 700 to 800°C, LHV decreases first, but finally it increases to its maximum value 8.56 MJ/Nm^3. They stated that by going to higher temperatures, from 800 to 900°C, LHV value experiences lower amounts and finally reached to its minimum value at 900°C. In similar temperature range, (750-900°C). Li et al [39] investigated steam gasification of EFB in a fixed bed reactor. They declared that in higher temperatures hydrogen production enhanced until its maximum value at 900°C. They indicated that in this temperature enhancement process concentrations of H_2 and CO_2, increased while the concentration of CO and CH_4 decreased, as well. According to them, LHV decreased continuously at this temperature enhancement process. In another research which is carried out by Womgsiriamnuay et al. [29] on a fluidized bed, bamboo was used as feedstock and air as an agent. Their results followed different behavior during the enhancement of temperature. They showed that the concentrations H_2 and CO decreased but concentration of CO_2, increased,

while the temperature increased from 400°C to 600°C. On the other hand maximum LHV value and CCE occurred at 500°c. Table 2.1 **Appendix B** summarizes studies which have investigated the effect of temperature on gasification of various biomasses. It can be inferred from these Investigations that higher amount of H_2 gas and lower concentrations of heavy tars and char will be obtained if the temperature increases. Skoulou et al. [40] stated that enhancement of H_2 production and decreasing tar concentration is due to tar thermal cracking reaction.

All the results can be justified by Le Chatelier's principle, when the temperature increases. Table 2.2 **Appendix B**, shows all the reactions which take place during the gasification. According to the principle, temperature enhancement will push the reaction in a way that endothermic reactions improve and exothermic reactions occur in opposite direction as showed in table 2.2 **Appendix B.** So, reactions such as R2 (boudouard), R3 (water-gas), and R8 (steam-methane reforming) will improve. The result of improvement in R3 and R8 reactions is enhancement of H_2 production [39]. Increase in R8 reaction also increases the CO production and CH_4 consumption. As a result, CH_4 concentration will reduce [38]. It can be inferred from R2, R3, and R8 reactions that CO production will enhance if temperature increases [40]. In addition, exothermic reaction R1 (char partial combustion) affects CO production, negatively. When the temperature increases, the content of CO decreases. CCE enhancement is predictable while temperature increases. Because it pushes the R2 and R3 forward and more carbon and steam are converted [38]. To put it in nut shell, higher temperatures favors gas yield and H_2 production. On the other hand, temperature enhancement can't always increase the LHV as too high as it lowers the gas heating value [38].

2.3.3 Biomass particle size effect

The gasification process can be significantly impacted by the feedstock's particle size. In general, quicker gasification rates and better gas yields are caused by smaller particle sizes.

The surface area to volume ratio is larger when the feedstock particles are smaller. This indicates that there is a larger surface area accessible for the gasification agents—such as steam or oxygen—to react on. As a result, the reaction can go forward more quickly and produce more gas. However, utilizing lower particle sizes could have some negative effects as well. For instance, smaller particles may be more challenging to manage and more likely to clump together or agglomerate, which could lower gasification efficiency. However, bigger feedstock particle sizes might be the outcome.

Mohammed et al. [15] has investigated biomass particle sizes from 0.3 to 1.0 mm and stated that the particles in this range doesn't cause blockage in feeder. They have shown that tar and char yields, have been increased and total gas yield has been decreased by particle size enhancement. They also declared that smaller particles generate lower CO_2 and higher amounts of CO and CH_4 in comparison with larger particle size. Furthermore, H_2 production has been remained approximately the same in every particle sizes in this range and even smaller than 0.3 mm. According to them, the highest and optimized value for LHV were obtained when the particle size was between 0.3 and 0.5 mm.

Li et al. [39] investigated particles with size between 0.15 and 5mm. They reported that H_2 and total gas yield decreased by increasing the size of particles. They stated that more H_2 and CO_2 and less CH_4 and CO was produced while using smaller particles. In addition, they

reported that maximum value of LHV took place when the particle had its largest value. Lv et al. [38] investigated another research and stated that smaller particles cause higher amounts of CO and CH_4 production and less CO concentrations. The smallest particle size in this research resulted in maximum value of gas yield, LHV and, CCE amounts. Luo et al. [27] investigated particle sizes between 0.075 and 1.2 mm. They stated that H_2 production and CCE decreased and char and tar amount increased, when particle size of feedstock enhanced.

Yan et al. [42] utilized particle size between 0.15 and 3mm and reported different behaviors. They conducted the experiment at 850°C and stated various behavior during the particle size enhancement. They stated that when the particle size increased from 0.15 to 0.45 mm, gas yield increased but with further increase in particle size, it declined, slightly. The maximum value of gas yield occurred when the particle size was between 0.45 and 0.9 mm. They declared that particle size had no effect on the molar fraction of syngas. Table 2.3 **Appendix B** summarizes investigations according to the particle size and its effect on syngas production.

To cut a long story short, less char and tar, more H_2 production and higher gas yields are the results of using smaller parts. As larger particles causes more temperature gradients within the particles and particle surface, then, particle's surface experiences higher temperatures in comparison with its core. Therefore, char and liquid yields increase and gas yield decreases [15]. On the other hand, more surface area is available in smaller feedstock particle sizes. So, they can absorb heat rates, faster. In addition, this high heat absorption can produce higher concentrations of light gases and lower amounts of char and condensate.

Another advantage of smaller particles and apparently higher surface area is enhancement in contacts between biomass and steam which provides more reaction rates and adequate gasification reactions [39]. It has been stated in same investigations that particle size has the lowest impact factor among other parameters. But, it's necessary to indicate that more reaction time is required when particle sizes are larger However, this effect will have less significance when operation temperature is high.

2.3.4 Gasification agent effect

The gasification agent, which can be oxygen, air, steam, or a mixture of these, plays a crucial role in the gasification process. It affects the reaction kinetics, product composition, and gas yield. The choice of gasification agent depends on the desired product and the feedstock being used.

One of the most important effects of the gasification agent is to control the amount of oxygen available during the reaction. In gasification, the feedstock is partially oxidized to produce a synthesis gas, or syngas, which contains a mixture of carbon monoxide (CO) and hydrogen (H_2) as well as other gases such as methane (CH_4) and carbon dioxide (CO_2). The amount of oxygen present during gasification affects the composition of the syngas. If the gasification agent is pure oxygen, the syngas produced will have a higher CO concentration than if air or steam is used.

Another effect of the gasification agent is to influence the reaction rate. For example, steam can enhance the gasification rate because it reacts with the carbon to form CO and H_2, which are the primary components of syngas. Air, on the other hand, can cause slower gasification rates because it dilutes the reaction with nitrogen and other inert gases.

The gasification agent also affects the operating temperature of the gasifier. In general, the higher the temperature, the more complete the gasification reaction, and the higher the gas yield. However, excessive temperatures can cause thermal cracking, which can produce undesired products such as tars and other complex hydrocarbons. The gasification agent can be used to control the operating temperature and minimize the formation of these byproducts.

Overall, the choice of gasification agent is an important consideration in gasification. Different agents can produce different gas compositions, affect the reaction rate and operating temperature, and influence the formation of byproducts. The choice of gasification agent depends on the specific feedstock and the desired product.

Recently, there have been various biogas gasification with different agents including steam, oxygen, air, or mixtures of these components. Table 4 summarizes some of these investigations.

Lv et al. [43] compared air and steam as agent in gasification process and stated that steam agent produces more H_2 than air as an agent. They used pine wood as feedstock within a down draft fixed bed gasifier. The hydrogen yield for air and steam as an agent were 0.33 and 0.49 m^3, respectively. They reported concentration of H_2 and CO, 63.27 - 72.56% for biomass with oxygen/steam gasification, but these values for biomass/air gasification were 52.19 - 63.31%, respectively.

Wongsiriamnuay et al. [29] observed that the value of CCE and total gas yield is higher for air/steam agent in comparison with dry air. In addition, they reported that H_2 and CO concentrations in product gas were higher, also. Li et al. [39] observed that enhancement in the steam/biomass ratio (SB) from SB = 0 (dry air) to 1.33 resulted in maximum amount of H_2 and gas yields. By increasing SB to higher values (SB= 2.67), the H_2 and gas yield experienced decreasing behavior.

Chang et al. [44] observed the same trend when they investigated the α-cellulose effect on gasification at 800°C. They repeated the process of gasification for an equivalent ratio of 0.27 with various SBs from 0 to 1.5. From 0 to 1 SB values, results presented that H_2 production, enhanced from 13.5 vol% to 18.56 vol%. Further enhancement caused decrease in H_2 concentration which can be concluded that maximum value for H_2 and total gas belongs to SB value of one. They stated that the lowest value for LHV was occurred in SB value of 1.5.

Mohammed et al. [15] studied the equivalent ratio (ER) in gasification process by increasing ER From 0.15 to 0.35. This resulted in reductions of char from 13.65 to 2.12 wt% and tar amount, from 9.830 2.82 wt%. They also claimed that LHV of produced gas declined to 12.35 from 15.38 MJ/m^3. They recorded maximum value for total gas production and H_2 concentration at ER of 0.35 and 0.25, respectively. They reported increase in CO_2 production and decline in CH_4 and CO generation. Skoulou et al. [40] investigated ER effect on gasification process and reported that by increase in ER value from 0.14 to 0.42, there was a decline in char and tar concentration. On the other hand, total amount of gas yield maximized at ER value of 0.21 and decreased when ER reached to 0.42. Both LHV and H_2 values, reached to their maximum and minimum values at ER value of 0.21 and 0.42, respectively. CH_4 and CO followed similar behavior when ER value increased. Those gases reduced as ER values increased. In contrary, CO_2 concentration first showed enhancement but with further increase in ER value, CO_2 amount experience redaction. This maximum production took place at ER value of 0.42.

Chang et al. [44] reported higher amounts of H_2 yields when steam was existed in agent gases. According to them, higher SB values concluded with higher values for total gas and H_2 yields. This SB ratio enhancement, favors steam consuming reactions (R3, R7 and R8), which contributed in enhancement of H_2, CO_2 and total gas concentration increase [39]. In addition, CO and CH_4 concentration will decline [45]. Increasing steam in agent has its own disadvantage and one of them is that steam is absorbs energy from the system and has negative effect on energy production [45]. Introducing extra steam may result in reduction of the reaction temperature which will cause in low quality gas production, consequently [39].

According to above mentioned investigations, higher ERs result in lower amounts of char, tar, LHV, CO and more CO_2, concentrations. Skoulou et al. [46] noticed a contradictory behavior for ER increase in gasification processes.

When ER increases gas quality reduces. The reason is that by oxidation reaction which results in CO_2 production enhancement, and simultaneously, lower production of combustible gases. Too high ER values contributes in less H_2 concentrations (R6), and CO_2 concentration enhancement in produced gases (R5) [15]. All in all, high ER values lead to exothermic oxidation reactions which, simultaneously, introduce further heat for gasification, and will optimize the quality of product, partially [46]. However, in very small values of ER, temperature will be low and consequently won't be appropriate for reactions in gasification process [43]. In overall, ER value affects quality of produced gas in gasification process [43].

2.3.5 Bed Material Effect

The bed material in a gasifier plays an important role in the gasification process, affecting the gasification efficiency, product gas quality, and the lifetime of the gasifier.

The bed material provides a surface for the gasification reactions to occur and helps to maintain the required temperature inside the gasifier. It also acts as a filter for particulate matter and helps to promote the mixing of the gasification agents with the feedstock.

One important effect of the bed material is its ability to store and release heat. Some common bed materials, such as sand, can store heat and release it slowly over time, which helps to maintain the temperature inside the gasifier. Other materials, such as ceramic beads or coke, have low heat storage capacity and can allow for more rapid temperature changes. The choice of bed material depends on the gasification process requirements, such as temperature, residence time, and desired product.

The bed material also affects the gasification reactions by altering the gasification agent's composition. For example, in fluidized bed gasification, the bed material is typically an inert material such as sand or alumina, which does not react with the gasification agents. However, in some cases, the bed material may be a catalyst that promotes the gasification reaction and can increase the gas yield or improve the product gas quality.

The bed material can also impact the lifetime of the gasifier. Over time, the bed material can degrade due to mechanical wear, erosion, or chemical reactions with the feedstock or gasification agents. This can lead to the formation of ash or other undesirable byproducts, which can reduce gasifier efficiency and require more frequent maintenance.

Overall, the choice of bed material is an important consideration in gasification. Different materials can have significant effects on the gasification efficiency, product gas quality, and

the lifetime of the gasifier. The selection of the bed material depends on the gasification process requirements, such as temperature, residence time, and desired product.

Some investigations has studied effect of sorbent for absorbing the CO_2, during the gasification process. Mahishi and Goswami [48] investigated gasification process of pine bark for biomass molar ratio of 1 relative to CaO. This molar ratio was defined by thermodynamic calculations. First, they reported that at 600°c of gasification process and absence of sorbent, total gas yield, CO_2, and H_2 gas composition and CCE were 0.87 m3/kg, 27.7 vol%, 60 vol% and 20.3%, respectively. While these values in the presence of sorbent increased for total gas yield, H_2 and CCS to 1.92 m3/kg, 64.5 vol% and 65.6%, respectively and CO_2, concentration decreased to the 26.8 vol%. It can be inferred that CaO played the role as sorbent but also as catalyst. Hanaoka et al. [49] used the same sorbent for oak bark at 650°C with molar ratio of CaO to the carbon in the biomass ([Ca]/[C]) values 0, 1, 2, and 4. For the values 1, 2, and 4, no CO_2 observed at the products. They reported the maximum value for hydrogen yield at ([Ca]/[C] = 2) which were 0.8 m³/kg.

The greatest challenge during the gasification process is tar formation which is a compound of condensable hydrogen and is the mixture of one to five aromatic ring compound connected to other oxygen-containing hydrocarbons and complex poly aromatic hydrocarbons PAH [50]. Some researchers investigated the effect of catalyst particles on improving the H_2 yield due to tar decomposition [51]. Shen et al. [51] stated that sand is the most widely used bed particle with very favorable mechanical properties which is particularly used in bubbling and circulating fluidized bed applications. It should be noted that sand never engages in chemical reactions with biomass. Moreover, a Ni catalyst and, limestone can play role both as a mechanical activity and resistance [51]. Wu et al [52], investigated nickel based catalysts which are cheaper in comparison with other noble metal catalysts. They have reported that group of catalysts are the most effective and promising catalyst.

Valliyappan et al. [53] used glycerol in a FIB gasifier by using Ni/Al_2O_3 as catalyst. They reported that by increasing the catalyst amount from 0 to 0.8 wt%, H_2 concentration enhanced from 55.9 to 68.3 mol%. Wongsiriamunuay et al. [29] investigated gasification process with calcined dolomite as catalyst with various catalyst to biomass ratios (CB) of 0:1, 1:1, and 1.5:1. By temperature and CB enhancement they reported increase in H_2 and CO yields and in contrary decline in CH_4 and CO_2. In addition, they stated that LHV and CCE improved due to enhancement of H_2 and CO concentration. They stated for reforming reaction enhancement due to catalyst presence which reduced the amount of produced tar. Li et al. [39] studied tri-metallic catalyst and compared the gasification results for calcined dolomite with tri-metallic catalyst and without catalyst processes. Table 2.5 **Appendix B** summarizes those results in the presence and absence of catalysts in the gasification process.

These investigations show that presence of catalyst and CaO can reduce the tar formation and cause enhancement in total gas and H_2 production and CCE. It has been stated by all researchers that amount of CO_2 absorption is tightly a function of partial pressure of CO_2 in generated stream at a constant gasification temperature. When the gasification temperature exceeds the equilibrium temperature corresponding to the CO2 partial pressure, CO2 is absorbed, converting the sorbent to CaCO3 in the process.

However, when above mentioned temperature is lower than the gasification temperature, $CaCO_3$ is released to form the original CaO. CO shift reaction (R7) states that in the presence of CaO, reaction tends to move RHS of the reaction to produce more H_2. So, as stated earlier,

CaO plays either a sorbent role or a catalyst role [48]. As the tars and hydrocarbons are reformed in the presence of calcium oxide, additional hydrogen is produced, calcium oxide serves the dual duty of a sorbent and a catalyst [39]. Catalysts cause tar cracking and increase tar reforming reactions, which results in H_2 concentration enhancement.

In general, hydrogen-rich syngas is often beneficial to produce biomethane after coupling biological systems such as anaerobic digestion. Otherwise, additional hydrogen need to be injected inside the reactor for the complete conversion of CO and CO2 from syngas.

2.4 Anaerobic digestion process parameters

Anaerobic digestion utilizes organic waste as feedstock, such as municipal solid waste, residual waste from agriculture and forest for biogas production [54]. A mixture of anaerobic microbes degrades the organic material and produces biogas, a mixture of carbon dioxide, methane, hydrogen sulfide, ammonium, etc[54]. Not limited to organic waste, this technology can be used for methane production using gaseous products such as CO2, H2 and CO [55]. Recently biomethane production from syngas has been widely researched in lab-scale reactors [56]. The AD has been coupled with syngas derived from the gasifier to produce the natural gas grid-quality biomethane (>97% CH4) [35]. The natural gas quality biomethane (>97% CH4) production from biogas or syngas is called gas upgrading [55], [57]. Gas upgrading is essential for the application of biogas, such as transportation fuel and gas grid injection [57]. The recent research result reported that CO2 species dominate in syngas derived from the gasifier; therefore, additional hydrogen injection could benefit from converting entire CO2 ad CO into biomethane (>97% CH4) [35]. Thus, this thesis investigated the techno-economical analysis of the coupling of the gasifier and anaerobic digestion for biomethane production..

In this section, the most important parameters affecting the anaerobic digestion process will be discussed according to investigations carried out on these parameters. Parameters are temperature, PH, C/N/P ratio, organic loading rate, and organic toxicants [61].

2.4.1 Temperature Effect

Increase in temperature will result in microbial metabolism rise due to biogas production enhancement. For instance, biogas generation rate for food waste in thermophilic circumstances (40-65°c) is higher than mesophilic conditions (25-40°c) which is higher than psychrophilic circumstances (10-25°c). Meanwhile, practically, AD takes place at two last conditions and is the least probable to occur at mesophilic circumstances. Because, bacterial growth is limited in that temperature level [62]. Sun et al. [63] stated that the most appropriate temperature for psychrophilic methanogens is 15 °C. According to table 2.6, at 15°C for rice digestion, one need 26 days period and 15 days is the peak time (the highest bio gas production rate) which is achieved by Xiao et al. [64]. This temperature is 35°c for mesophilic conditions. According to Xiao et al. [64] peak time for biogas generation would be 3 days, which is shown in table 2.6. They reported this time 3 days for straw or food waste and 55°c was the suitable temperature in thermophilic conditions, according to Mao et al. [65]. AD process, not only in mesophilic but also in thermophilic conditions, is sensitive to temperature variations and should be kept around 3°C [66]. If temperature is not controlled in that temperature the biogas production will reduce and will be fully stopped around 5°C. According to Walter et al. [67] thermophilic methanogens can withstand temperature fluctuations within 5°C/h.

2.4.2 Effect of pH

Hydrolytic bacteria prefers neutral environment. Therefore, pH of environment should be neutral where hydrolysis of organic substrates will take place. So, pH should be maintained within a specific interval (5-7) to keep normal functionality of microorganisms [68]. Excessive acidic or alkaline conditions will reduce the hydrolytic efficiency of them. In addition, this pH range is also dominant for acidogenic microbes. For instance, Yang et al. [69] investigated the digestion of rice and meat wastes in production of acetate and ethanol. They reported 20-39% production enhancement of acetate by maintaining the pH in the 4.5-6 interval. These results has been depicted at table 2.6 [70].

Dareioti et al. [71] reported that for pH values over 5.5 or less than 9.5, production of propionate can be declined into 42.5%. Although acetogenesis can take place either in acidic or alkaline environments, acetogenesis microbes life can be threatened in extra acidic or alkaline pH values [72], [73]. For instance, acetogenesis bacteria can die at 5-8 pH values [61]. Lee et al. [74] reported that all acetogenic bacteria groups produced acetic acid during the digestion of corn and wheat, properly, when the pH value was maintained in a 5.8-6.8 interval. In addition, this values of pH is essential for methanogenesis bacteria. According to table 2.6, Bassani et al. [75] and Fu et al. [76] have stated that if pH value is less than 6.3 or more than 7.8, the methanogenesis rate will be declined to 56.9% and reported that those microorganisms aren't able to survive outside the 7-8 pH value range.

2.4.3 Effect of C/N and C/P ratios

The other parameter that can affect the AD performance is carbon-nitrogen-phosphorus ratio. Food wastes, manure, and plant residues are the main source of carbon [77]. Organic solid wastes (OSWs) which are mainly protein and urea are the primary sources of nitrogen [78]. Organic Orthophosphate-phosphorus (PO_4-P) and polyphosphates are the origination point of Phosphorus. Either P or N are vital and necessary not only for growth and synthesis of microbial cells but also for pH stabilization under the alkaline conditions [79].

When the C/N ratio is less than enough, the extra total ammonia nitrogen (TAN), harms the process of methanogenesis, because of three reasons: 1) It inactivates methanogens enzymatic activities. 2) It disturbs proton equilibrium. 3) It causes lack of potassium ions (k^+) in cells with extra TAN [61]. According to Huang et al. [80], biogas production efficiency will decline 65% when TAN value is 4 kg/m³. But, it will increase when TAN value increases from 0.05 to 0.4 kg/m³. Finally, they propose the 25-30 value of C/N ratio as the most optimized range for AD process. On the other hand, anaerobic bacteria needs external P sources because they are unable to synthesize organic P by themselves. Investigations have concluded that for carbon to phosphorus ratio (C/P) value of 157 the production efficiency of VFAs in acidogenesis process will be 17.5% [81]. As a result, researches have shown that the most suitable C/P ratio is 150 and C:N:P value for AD process is 200-300:5:1 [61]

2.4.4 Organic loading rate (OLR) effect

There are two ways for reaching higher QLRs. First, by reducing the hydraulic retention time (HRT). Second, by substrate concentration enhancement. Although, OLRs are normally in a range of 5-40 kg COD/ m² day, it can even reach to higher values in AD process [82]. As can

be inferred from table 2.6, biogas production efficiency can be increased by 43% while ORL enhances from 11.12 to 20.26 kg COD/m² day during AD process of sugar cane straw [83]. This efficiency even has increased to 57% by further enhancement of OLR into 24.4 kg COD/m² day. However, extra food for microorganisms will result in enhancement of pressure at AD process. The reason is that acid bacteria cannot digest all kinds of loaded materials and will concluded with an extra VFAs [84]. During the cow manure digestion, ORL has been increased from 18.44 to 37.9 kg COD/m² day which resulted in biogas production efficiency reduction by 39% [85]. Baek et al. [86] reported that when ORL is more than 51.8 kg COD/m² day, the risk of death is more probable to occur for anaerobic microbes.

2.4.5 Effect of organic toxicants

Several organic toxins, are toxic to methanogenesis process such as long chain fatty acids (LCFA). The reason is that LCFAs reduce the surface tension of microbial cell membrane and cause permeability. As can be inferred from table 2.6, the efficiency of biogas production has been reduced 40.2% by increasing lauric acid and plamitic acid. When acid concentration exceeds 2.7 kg/m², it will kill the methanogens [87]. The other group, is pharmaceuticals such as carbamazepine, sulfamethoxazole, and propranolol hydrochloride, which are harmful for acetoclastic methanogens [88]. These pharmaceuticals can prevent growth and generation of methanogens or can diminish their enzyme activities which both can result in failure of AD system. For instance, when the value of carbamazepine amount is higher than 0.23 kg/m³, methanogenesis will cease, and will result in acetate accumulation [89]. The other toxicant is cyanide which is generated during the sweet potato cultivation process. It inhibits the enzymatic activities, and finally decreases synthesis and division of cell and normal processes of methanogens. In a specific time interval, cyanide can exacerbate methanogen toxication. Efficiency of biogas production has been reduced 51.5% and time has increased from 4 h AD to 10 h AD, when the cyanide is 2.42 X10 kg/m³ [90].

2.4.6 Metallic ions effect

Microbial activity and growth can be stimulated by even low amounts of light metals such as aluminum (Al^{3+}), magnesium (Mg^{2+}), potassium (K^+), and sodium (Na^+). External metallic ions addition, is essential for anaerobic bacteria because they are not able to produce all of their needs to these metal ions [91]. But further increase in metallic ions will inhibit the functionality and will cause toxication. For instance, higher values than 2×10^{-3} kg Al^{3+} /m³ will result in reduction of acid-forming and Methanogen's activity due to adherence of Al^{+3} to the cell membrane and prevents the transportation processes [92]. The most suitable amount of Al^{3+}, Mg^{2+}, K^+, and Na^+ are less than 0.428, 0.78, 0.476, and 0.512 kg/m³, respectively.

Zhang et al. [93] has stated that when the concentration of potassium ion is 0.387 kg/m³, biogas production rate will be the maximum value by 37.5%. On the other hand, digestion efficiency can be reduced from 89.2% to 61.7% by further enhancement to 0.597 kg/m³ of potassium.

In specific concentrations, heavy metal ions are also able to improve anaerobic bacteria functionality [94]. Heavy metals which can be found in OSW's (organic solid wastes) are iron (Fe^{3+}) copper (Cu^{2+}) chromium (Cr^{3+}) and cobalt (CO^{2+}) ions. All in all, heavy metallic ions are supposed to be toxic for AD process [95]. The main reason for being toxic is reaction with microbial proteins and enzymes. These reactions result in reduction of methanogens and acid-

formers' activities. For example, Cu^{2+} will keep the normal function of AD when its concentration is under 5.88×10-4 kg/m³ but biogas production efficiency will be reduced by 12.9% when the concentration reaches to 4.7×10^{-3} kg/m³ [96]. So, the total value of heavy metallic ion in AD process has been proposed less than 1.5×10^{-3} kg/m³ [97].

There is growing interest in coupling anaerobic digestion and gasifier to utilize waste as resources in the circular economy [55], [56]. The lab scale experiment has shown promising results; however, limited or no investigation was dedicated to evaluating the techno-economic evaluation of the coupling scenario for biomethane production. Therefore, a systematic techno-evaluation was performed to fill the gap from the recent state of the art.

There have been studies on coupling gasification and anaerobic digestion processes to increase the amount of biomethane production. This process is called integrated gasification and anaerobic digestion (IGAD), and it involves gasifying the solid feedstocks to produce syngas, which is then fed to an anaerobic digestion reactor along with liquid or semi-solid feedstocks.

The gasification process converts the solid feedstocks into syngas, which contains carbon monoxide, hydrogen, and other gases. The syngas is then fed to an anaerobic digestion reactor along with liquid or semi-solid feedstocks, where it is converted into methane and carbon dioxide by the anaerobic microorganisms.

The main advantage of IGAD is that it can increase the amount of biomethane produced from a given feedstock, by utilizing the syngas produced from the gasification process. This can improve the overall efficiency of the process, as well as the economic viability of the project.

Several studies [58], [59] have shown that IGAD can be an effective process for producing biomethane from various feedstocks, including agricultural residues, municipal solid waste, and sewage sludge. However, IGAD is a complex process that requires careful integration of the gasification and anaerobic digestion systems, as well as careful selection of the feedstocks and operating conditions.

One recent investigation focused on the feasibility of using IGAD for biomethane production from food waste [58]. The researchers used process simulation and techno-economic analysis to evaluate the performance and economic viability of the IGAD process.

The results showed that IGAD was a promising process for producing biomethane from food waste, with a biomethane production potential of 427 m^3/tonne of food waste. The study also showed that the economic viability of the process was dependent on several factors, including the feedstock cost, electricity price, and biomethane price.

Overall, the study demonstrated the potential of IGAD as a sustainable and economically viable process for biomethane production from food waste, which could contribute to reducing greenhouse gas emissions and promoting a circular economy.

2.Literature review

3 System overview

 The current chapter introduces the method for economic and simulation process of combining syngas fermentation and hydrothermal gasification to produce biomethane. Through the process design, the syngas fermentation is coupled with gasification process. The study intends to assess the economic viability of the three scenarios for the birch wood waste gasification to produce methane.

3.1 System arrangement

The integrated processes of the methane biorefinery are shown schematically in Figure 1. The flowchart employed in this study was used to evaluate the following three scenarios. The choice of three scenarios was made based on the results derived from the literature. The coupling of the gasifier with anaerobic digestion resulted in different methane content. It has been demonstrated that hydrogen is one of the limited gas factors in achieving biomethane (>97% CH_4) production. Thus three scenarios were created based on understanding from literature as i) utilize syngas composition as derived from gasifier ii) utilize steam to produce hydrogen-rich syngas, and iii) utilize additional hydrogen for complete conversion of CO and CO_2 into CH_4

(i) Scenario 1 (generation of biomethane without any additional hydrogen): The most straightforward setting serves as the starting point to conduct extra comparisons. Hydrothermal gasification and syngas fermentation are the components of Scenario 1 (Fig. 1a).

(ii) Scenario 2 (generation of biomethane plus adding extra hydrogen): In this scenario, steam is injected as hydrogen source during syngas fermentation (Fig. 1b).

(iii) Scenario 3 (biomethanation plus pure hydrogen injection): The technique used in this case is comparable to that in scenario 2, with the exception that hydrogen is added to fermentation stage by a PEM electrolyzer unit (Fig. 1c). It should be emphasized that the syngas fermentation reactor's dimensions are equal to those in scenario 2 and 1.

In general, the study plans to evaluate different conceptual designs for biomethane production in order to enable economic evaluation. This will be done through extensive techno-economic and sensitivity analysis. Key economic indicators and the circumstances in which they can promote the commercialization process could also be gained through this investigation.

(a)

(b)

(c)
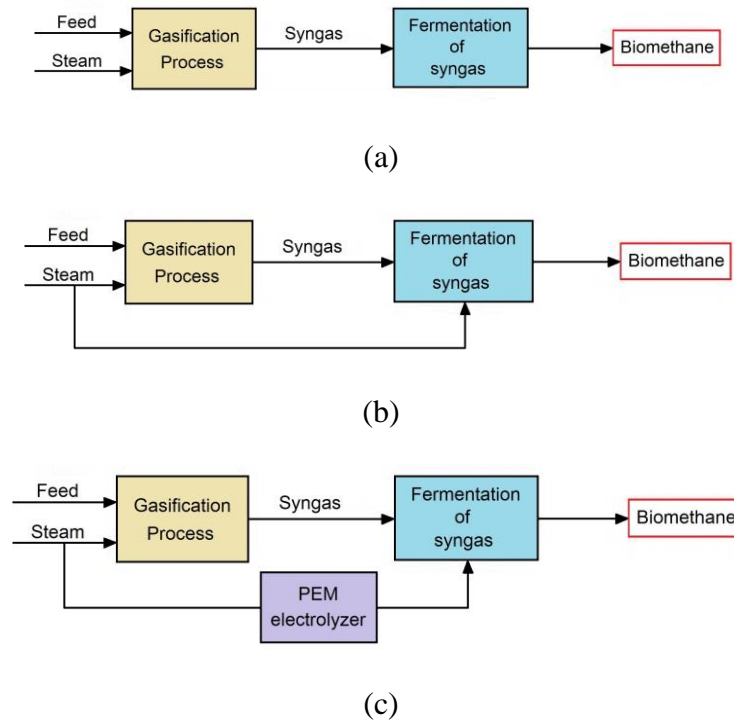
Figure 3.1 Biomethane production through (a) scenario 1, (b) scenario 2 and (c) scenario 3.

3.2 Process simulation

Fig. 3.2 depicts the proposed design for a biomethane production process flow diagram. The system's mass flow and energy flow were measured by the conceptual design. For this purpose, Aspen Plus v12 (AspenTech, Bedford, USA) was used to model a number of unit operations, including hydrothermal gasification, syngas fermentation, water separation, and cooling. The facility is designed to handle about 896 metric tons of birch wood waste, annually. It should be mentioned that Aspen Plus has a powerful database of physical properties that was employed for calculations.

Figure 3.2 Biomethane production simulation flow sheet (Senario 3)

3.2.1 Process description

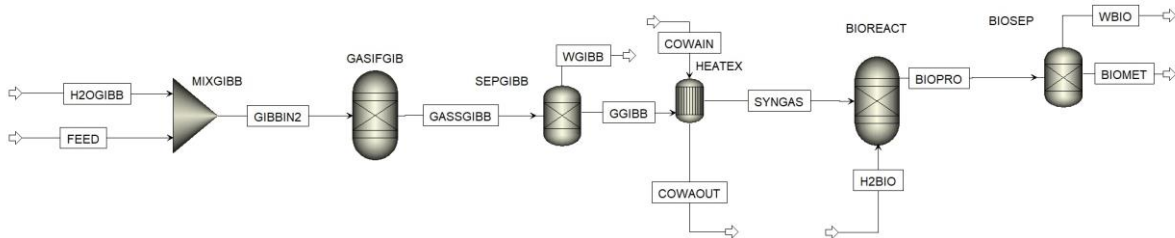Before entering the gasification reactor, the feed stream is mixed with the gasification agent, in this case steam, in a mixer (MIXGIBB). Water is eliminated from the reactor's output via separator (SEPGIBB) in order to accommodate for a dry analysis of the gas composition.

The computation is carried out at reactor temperature (850°C) in the Gibbs reactor (GASIFGIB). The parameters were developed using the same dimensions as those used in the work by Eikeland et al. [98]. After gasification process, in order to reach temperatures in which bacterizes can be active, the temperature is reduced to 50°C by means of heat-exchanger (HEATEX).

After temperature adjustment syngas is introduced to anaerobic digestion reactor (BIOREACT) for fermentation process. According to scenario1 no water, scenario2 water, scenario3 hydrogen, is added to this reactor in this step. In the final step, water is removed from the final product by separator (BIOSEP).

3.2.2 Composition of biomass

Table 3.1 displays the characteristics of the birch wood utilized as feedstock in this study. To facilitate experimental confirmation, the characteristics of the feed produced by Eikeland et al. [98] were utilized. In the next parts, unit operations were carefully detailed.

Table 3.1 Birch wood composition used in Aspen Plus simulation [98].

Gas	Mass fraction of mixed sub-streams mass	Mass fraction of CISolid sub-streams mass
CH_4	0.0969	0
CO	0.5307	0
CO_2	0.1365	0
H_2	0.0125	0
C	0	1
H_2O	0.1509	0
C_3H_6	0.0725	0

3.2.3 Components

Aspen Plus offers a large variety of components that can be utilized to simulate different chemical processes. The followings are a few of the common elements offered by Aspen Plus.The chemical elements hydrogen (H_2), carbon (C), nitrogen (N_2), oxygen (O_2), sulfur (S), and many others are listed in Aspen Plus' extensive list and can be used to simulate chemical

reactions.Aspen Plus contains a variety of inorganic substances, including water (H_2O), carbon dioxide (CO_2), ammonia (NH_3), sulfur dioxide (SO_2), and a lot more.Alkanes, alkenes, alkynes, alcohols, ketones, aldehydes, esters, ethers, aromatics, and many other forms of organic compounds are available in Aspen Plus' extensive database of organic compounds.

Ionic compounds: Aspen Plus contains a variety of ionic compounds that can be used to simulate ionic reactions, electrochemical processes, and other related systems. These compounds include salts, acids, bases, and other electrolytes.

Solid-state reactions, heat transfer, and other related processes can be modeled using solid-state compounds including minerals, catalysts, and other solid materials that are included in Aspen Plus.

Custom components: Aspen Plus enables users to create their own custom components with user-specified parameters. These components can be used to model proprietary chemicals or special chemical systems that are not represented in the database of standard components.

These are just a few illustrations of the common parts that Aspen Plus offers. Users can also add their own custom components as needed to the software's extensive library of parts, which can be used to model a variety of chemical systems and processes.

All of the components were correctly defined at the start of the simulation. The simulation's modelled components are listed in Table 3.2. Carbon was classified as solid components due to the uncertainty surrounding its exact chemical compositions. Other components are conventional throughout the simulation.

Table 3.2 Components used in Aspen Plus simulation [98].

Component ID	Type	Component name	Formula
CH_4	Conventional	METHANE	CH_4
CO	Conventional	CARBON-MONOXIDE	CO
CO_2	Conventional	CARBON-DIOXIDE	CO_2
H_2	Conventional	HYDROGEN	H_2
C	Solid	CARBON-GRAPHITE	C
H_2O	Conventional	WATER	H_2O
PROPY-01	Conventional	PROPYLENE	C3H6

3.2.4 Properties

Aspen Plus utilizes the MIXCISLD stream class, and Peng-Robinson (PR) cubic equation of state with Boston-Mathias for property method which is called PR-BM function. For applications in petrochemical, refinery, and gas processing, the PR-BM property approach is advised. For nonpolar or barely polar mixtures, the PR-BM property technique is employed. Hydrocarbons and light gases like carbon dioxide, hydrogen sulfide, and hydrogen are a few examples [99].

3.2.5 Stream properties

Table 3.3 is a list of the precise specs for the biomass, water, and hydrogen. It should be notified that the streams for water and hydrogen are for second and third scenario, respectively.
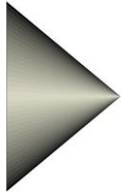
Table 3.3 Stream properties used in Aspen Plus simulation

Stream	Component	Temperature	Pressure	Mass flow rate	Engaged Scenarios
FEED	Table 3.1	0 °C	1 bar	112 kg/h	Sc1, Sc2, Sc3
H2OGIBB	H_2O	0 °C	1 bar	224 kg/h	Sc1, Sc2, Sc3
COWIN	H_2O	20 °C	1 bar	5000 kg/h	Sc1, Sc2, Sc3
H2BIO	H_2O	50 °C	1 bar	112 kg/h	Sc2
H2OBIO	H_2	0 °C	1 bar	10 kg/h	Sc3

3.2.6 Block properties

The next step (after defining the stream properties), is determining the properties for the blocks in the simulation. Table 3.4 shows the properties of blocks which is utilized in Aspen Plus simulation. It should be noted that these properties are equal for all scenarios.

Table 3.4 Block properties used in Aspen Plus simulation

Block name	Schematics	Temp	Press	description
MIXGIBB		NA	NA	Engineers and scientists can model the behavior of complicated mixtures and anticipate their attributes in a variety of process simulation scenarios using the Mixture block in Aspen Plus, which is an adaptable tool. In this simulation it has been utilized to mix the feed and water before entering the gasifier.
GASIFGIB		850 °C	1 bar	The Gibbs reactor in Aspen Plus is a type of reactor model used to simulate chemical reactions that occur under constant Gibbs free energy conditions. It is a specialized reactor model that allows for the prediction of the equilibrium composition of a reacting system based on the minimization of Gibbs free energy.
BIOREACT		37 °C		

SEPGIBB		NA	NA	The Component Separator in Aspen Plus is a unit operation model used for separating or fractionating a mixture into its individual components based on their physical properties, such as boiling point, condensation point, or vapor pressure. It is used to separate water from syngas and biofuel to obtain a dry basis combination.
BIOSEP				
HEATEX		NA	NA	The HEATX block in Aspen Plus can also include additional features, such as fouling factors, pressure drops, and temperature approaches, to account for realistic operating conditions in industrial heat exchangers. It has been used to cool down the syngas before entering the biomethanation reactor.

3.2.7 Assumptions

The Aspen plus simulation was based on a number of assumptions:

i.   The system runs under steady state circumstances and is isothermal.

ii. Operation pressure is atmospheric pressure, disregard for pressure drops.

iii. Ash, Sulphur, chlorine, and nitrogen are ignored.

iv. Tar formation is not taken into account.

v.   Char contains only carbon.

vi. The gasifier's heat loss is disregarded.

3.3 Economic evaluation

The minimum selling price (MSP) of biomethane—the price at which costs and revenues are equal—was calculated using an economic model, and its effects on various variables, such as the price of utility and labor cost, were also examined. The most significant techno-economic characteristics were also determined, and the net present value (NPV) and payback period (PBP) were later estimated, using a thorough discounted cash flow analysis (DCFA). This technique was also utilized to compare the three different scenarios examined in this study in terms of their financial feasibility. All the above mentioned parameters will be explained in detail in following sections.

3.3.1 Equipment purchase cost

The first step in economic evaluation is defining the EPC. Equipment which are used in various scenarios and their numbers are represented in the table 3.5.

Table 3.5. Equipment number in different scenarios

No.	Equipment's Name	Scenario No. Number of Equ.		
		Sc.1	Sc.2	Sc.3
1	Bioreactor for syngas fermentation	1	1	1
2	Electrolysis unit	0	0	1
3	Flash separators	2	2	2
4	Heat exchangers	1	1	1
5	Gasification reactor	1	1	1
6	Pumps	1	2	2

Available costs in different references are related to different years and various capacities. In order to reach required purchase cost, two sets of formulation have been utilized to reach this aim.

For converting purchase cost from initial year to current year (2023), Chemical Engineering Plant Cost Index (CEPCI) has been used as follow [100].

$$Cost_{Current} = Cost_{Ref} \left(\frac{CEPCI_{Current}}{CEPCI_{Ref}} \right)$$
(3.1)

CEPCI for different years is provided in table 3.6 **Appendix C**.

Although other formulations can be employed for various process units, the six-tenths-factor rule is the most common one. Using this approach, the equipment cost is calculated by multiplying the known cost of the real process units by the exponentiated ratio of the equipment capacities. An exponent suggests that economic aspects of scaling can be taken into consideration. Because it is highly helpful to scale up or down to a new capacity in order to determine the impact of a plant size. Eq. (3.2) illustrates the six-tenths rule [100].

$$\frac{C_{Current}}{C_{Ref}} = \left(\frac{A_{Current}}{A_{Ref}} \right)^n$$
(3.2)

Where C is the equipment's cost (NOK) and A is equipment's capacity (m^2, kW, kg/h…) and n is defined in the table 3.7 **Appendix C**. It should be noted that for USD to NOK currency conversion 10 is multiplied.

3.3.2 Capital Expenditure

To evaluate the CAPEX, the bottom-up methodology was applied. According to this method, each cost element was calculated as a percentage of the equipment purchase cost (EPC) [104]. The direct and indirect costs were calculated as part of the CAPEX estimate, as indicated in Table 3.6.

Table 3.8. Formulation utilized in the estimation of CAPEX (References: [102], [103], and [104]).

	Cost component	Factors used for estimation
DC	Equipment purchase cost, EPC (bare module cost) (1)	100% of EPC
	Equipment installation cost (2)	40% of EPC
	Controls and instrumentation (3)	26% of EPC
	Piping and electrical systems (4)	41% of EPC
	Buildings (including services) (5)	10% of EPC
	Yard improvements (6)	12% of EPC
	Direct cost, DC (7)	(7) = (1) + (2) + (3) + (4) + (5) + (6)=2,29*EPC
IC	Indirect cost, INC (8)	(8) = 21.9% of DC = 0,5015* EPC
FCI	Fixed capital investment, FCI (9)	(9) = DC + INC = 2,79 * EPC
	Working capital, WC (10)	15% of FCI
	Startup cost, SUC (11)	5% of FCI
	Capital expenditures, CAPEX	FCI + WC + SUC = 3,35*EPC

3.3.3 Operation Expenditure

The assumptions listed in Table 3.7 were also used to estimate OPEX, which includes both fixed and variable costs.

The cost of raw materials plus the price of utilities like cooling water and steam made up the variable operating cost (VOC) [103]. The fixed operating cost (FOC), on the other hand, was mostly tied to labor costs, employee social benefits, management, administrative costs, and maintenance and repair fees [104]. The number of workers employed for each 8-hour shift to run a fully automated process (N_{OL}), illustrated in Eq. (3.1) [105], was used for calculating the cost of labor.

$$N_{OL} = \sqrt{31.7P^2 + 0.23N_{NP} + 6.29} \qquad (3.1)$$

P is the number of processing stages involving the handling of particle solids, such as transportation and distribution, particulate size control, and particulate removal in this study P is considered 2 for first and second scenario and 3 for scenario 3. Where N_{OL} is the number of operators each shift. Compression, heating and cooling, mixing, and reactions are all included in the N_{np} that measures the number of non-particulate processing stages according to the Eq. (3.2) [105]. This parameter is equal to 3 for all scenarios.

$$N_{np} = \sum_{\substack{Reactors \\ Exchangers \\ Heaters \\ Towers}} Equipment \qquad (3.2)$$

An average yearly salary of 550,000 NOK was considered for each worker. Based on the correlation in Eq. (3.1), 12, 12, and 17 workers, were calculated to execute scenarios 1, 2, and 3 of the plant, respectively.

Table 3.9. Formulation utilized in the estimation of OPEX (References: [102], [103], and [104]).

	Parameter	Price
FOC	Supervision (1)	25% of labor cost
	Direct overhead (2)	50% of (labor + supervision cost)
	General overhead (3)	50% of (labor + supervision cost + direct overhead)
	Insurance and tax (4)	1% of FCI
	Maintenance labor and materials (5)	3% of FCI
	Additional expenses (e.g., marketing, logistics, operation services, etc.) (6)	1% of FCI
	Laboratory cost (7)	1% of Labor cost
	Financing working capital (8)	$0.1 \times$ WC
	Fixed operating cost (9)	(9) = (1) + (2) + (3) + (4) + (5) + (6) + (7) + (8)
VOC	Cooling water (10)	NOK 150/1000 m3
	Electricity (11)	NOK 1/KWh
	Total utilities cost (USD$) (13)	(13) = (10) + (11)
	Deionized water (12)	NOK 818/1000 m3 [1]
	Feed (14)	NOK 1 per kg [3]
	Total cost of raw materials (15)	(15) = (12) + (14)
	Variable operating costs (16)	(16) = (15) + (13)
	OPEX (17)	(17) = (9) + (16)

3.3.4 Economic theories and concepts

To comprehend and complete a techno-economic assessment, several economic principles should be defined.

The initial investment (C_0) and all subsequent cash flows (terms C_n, which do not account for the time value of money) are discounted back to their present value at a suitable hurdle rate (r), i.e., the sum of each net present cash flow (cash inflow - cash outflow) in year n over the entire number of years, N. This yields the net present value (NPV), which is the total of all present values, including the initial investment (C_0). So, in order to approve a project, cumulative or total NPV must be positive [100].

$$(total) \, NPV = C_0 + \sum_{n=1}^{N} NPV_n = C_0 + \sum_{n=1}^{N} \frac{C_n}{(1+r)^n} \tag{3.3}$$

The internal rate of return (IRR), which is based on incremental time-weighted cash flows, is the discount rate that results in the net present value zero. IRR must be greater than an acceptable rate in order to approve a proposal [100].

$$NPV = C_0 + \sum_{n=1}^{N} \frac{C_n}{(1 + IRR)^n} = 0 \qquad (3.4)$$

The payback period (PBP) is the amount of time that is needed to recoup the initial investment via business income. In other words, PBP shows how long it will take for the initial investment to be recovered from the generated cash flows, often stated in years when the working capital has been discounted.

The rate of return on investment (ROI) calculates the variance of revenues and investment in relative to each other, i.e., the efficiency of an investment in a business, after deducting costs.

The most prevalent analysis techniques for determining the viability of a project or its profitability are those based on PBP, ROI, NPV, or discounted cash flow upon full-life performance, however there are also alternatives.

It would be beneficial to evaluate projects with a brief lifespan because, according to definition, PBP does not take the project performance after the payback period into account. Similar to how ROI can be used to exclude investments with rates below a predetermined target ROI value and ignore the timing of cash flows. A discounted (or non-discounted) cash flow analysis, which is the recommended method to use and will be outlined below, can be employed to connect NPV and IRR.

The minimum selling price (break-even point) of biomethane can be determined using discounted cash flow analysis once the process's capital and operating costs have been determined. This cost is exactly what is necessary to recoup the initial investment without incurring any losses or gains, but at an interest rate determined by IRR. To complete this process, the following factors must be taken into account:

- Both the life time for performing the capital investment and the allocation for each year during that time period. Investment in land should be taken into account as well.

- The duration of both construction and operation.

- Taxation rate.

- The necessary IRR.

- The method of depreciating of assets and the life time for doing so; across the depreciation period, straight-line (or linear) depreciation is primarily used.

On the basis of a number of assumptions, illustrated in Table 3.7, the economic viability of a biomethane production plant with a birch wood processing capacity of 896 tons/year was evaluated.

Table 3.10 Assumptions for economic model estimation [106].

Parameters	Assumptions
Plant location	Norway
Currency used for economics analysis	NOK
Plant lifetime	20 years
Construction duration	1 years
Plant yearly operation	8000 h/year
Tax rate	30 %
Depreciation method	Straight line
Depreciation time	15 years
Salvage value	0 (used to offset decommissioning cost)
Base year	2023
Land cost	6% TIC
Feed utilization capacity	896 tons/year

Due to the harshness (high temperatures in gasifier) of the working conditions, lifetime should be greater than 15 years but not significantly longer (20 years might be a decent baseline). Depending on the timing and engineering, the depreciation may take between 10 and 15 years to complete. Similar to this, IRR should be 10% or more, as will be shown below [100].

The opening of a senior debt, which is borrowed money that a company must repay first, the funding costs, the interest, and other expenses must all be taken into account in a TEA. These expenses are typically financed by banks. A base-case analysis, or one that takes into account no borrowing, can be performed at 100% equity financing.

The formulas used in a spreadsheet to do the discounted cash flow analysis with 100% equity financing are provided in table 3.11 **Appendix C**.

In the economic calculation, it was also expected that the plant would not receive any further maintenance due to safety concerns, which resulted in a low salvage value (money from selling the used property at the end of its useful life). Additionally, the factory was built in a year and ran for nineteen years, 330 days a year, with a capacity factor of over 90%. 30% of income was taxed [100].

While the plant is being built, 50% of the TIC and land are paid. The most unfavorable cash flow is obtained right at this point because at the end of the first year, 55% of the total investment cost (TIC) (the remaining capital investment plus working capital) are paid, but nothing is created. The investment starts to be recovered at an IRR of 10% when production starts at the beginning of the second year. From this point on and through the last year, there are sales revenues (Oxygen in third scenario and biomethane) [100].

3.4 Sensitivity analysis

A sensitivity analysis for the MSP (Minimum Selling Price) of a biogas production plant is a tool used to evaluate the impact of changes in key parameters on the overall economic viability of the project. The analysis involves varying one or more input parameters while keeping all other parameters constant, in order to determine how sensitive the MSP is to changes in those parameters.

By conducting a sensitivity analysis for the MSP of a biogas production plant, project developers can identify the most important factors affecting the economic viability of the project, and develop strategies to mitigate potential risks or uncertainties. This can help ensure the long-term sustainability and profitability of the biogas production plant.

To make the proposed process configurations more reliable, consistent, and credible, sensitivity analysis was carried out on them. More significantly, the sensitivity analysis was employed to assess how different variables affected the MSP of biomethane. The factors with the most effects on the MSP of biomethane were also identified and ranked via sensitivity analysis. The price of utility cost of labor, cost of land, cost of biomass, and tax rate are a few of the variables. The variables were modified by giving them high and low values that were respectively +30% and -30% of their initial value.

3.5 Efficiency calculation

The energy efficiency of each process is determined using Eq. (3.1), which adds the energy content of the feed and the power consumption. It is based on the ratio of biomethane energy output (on an LHV basis). Since the other variables in Eq. (3.1) are written in terms of thermal energy, the latter was divided by 0.4 (efficiency of a typical thermal plant). LHV of feed is 15.02 MJ/kg on a dry basis.

$$\eta_{plant} = \frac{\dot{m}_{bio-CH_4}LHV_{bio-CH_4}}{\dot{m}_{feed}LHV_{feed} + Electricity\ consumed/0.4} \tag{3.1}$$

3.System overview

4 Results

 In this chapter simulation results has been validated by experimental results. In the next step, economical evaluation of design has been presented for different scenarios. Finally, sensitivity analysis has been illustrated for those three scenarios.

4.1 Gasification process validation

 A comparison between the simulation predicted gas yields and empirical values [107] is shown in Figure 4.1. This figure shows a slight deviation between gasification products through various temperatures. The outcomes demonstrated that hydrogen production (purple color) was enhanced at higher temperatures. In addition, findings showed that greater temperatures resulted in less carbon dioxide (green bars) production. Moreover, the results showed that carbon monoxide (red columns) production was boosted by higher temperatures. On the other hand, the findings of the simulation and the experimental data for methane (blue pillars) differ significantly from each other, in contrast to other components. It should be noted that more details has been provided in discussion chapter, specifically.
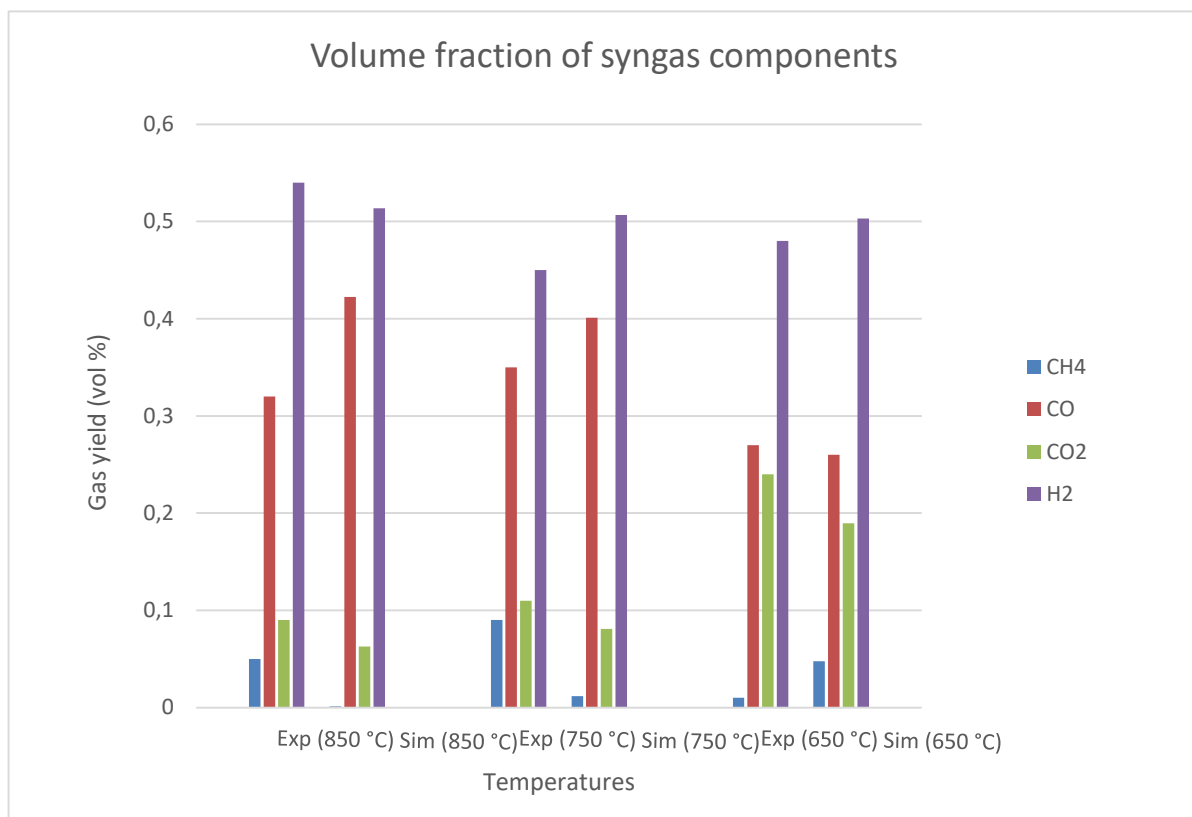

Figure 4.1 Comparing the empirical gas yield from [107] studies at various temperatures with the gas yield determined by Aspen Plus simulation.

4.2 Economical assessment

In this part of results the outputs of investigation and calculation on economical aspect of the financial concepts are represented for each of scenarios in different parts such EPC, OPEX, DCFL and, etc.

4.2.1 Equipment purchase cost

Fig. 4.2 shows the breakdown of the equipment acquisition costs for each scenario. Pumps and flash separators account for 29.4% and 35.2%, respectively, of the total equipment acquisition cost during Scenario 1 (left column). In scenario 2 (middle column), these items make up 45.4% and 27.2%, respectively, of the total cost of the equipment. Pumps, separators, and electrolyzers make up the majority of the equipment in scenario 3 (right column), accounting for 39.2%, 23.4%, and 13.7% of the total equipment purchase cost, respectively.
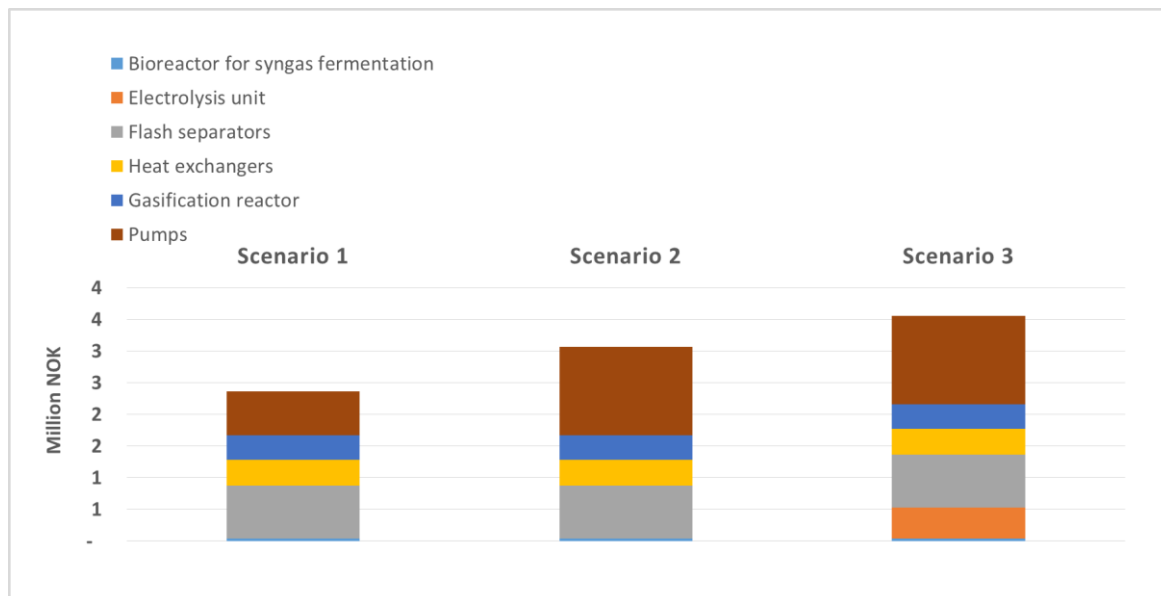

Figure 4.2 EPC breakdown for each scenario

It should be underlined that results related to this bar chart is represented in **Appendix D**.

Due to its straightforward design and lack of several processing units like those found in scenarios 2 and 3, Scenario 1 had the lowest overall equipment cost (2368057 NOK). As opposed to scenario 1, scenario 3's overall equipment cost was almost 1.5 times of scenario 1. Following is the sequence in which the overall cost of the EPC decreased: Scenario 3 (3552892 NOK) comes in ahead of scenario 2 (3064361 NOK) and scenario 1 (2368057 NOK).

4.2.2 Operational and capital expenditures

Fig. 4.3 displays the breakdown of OPEX for each scenario. It should be mentioned that the price of feed cost and the cost of the utilities were included in the variable operating cost

(VOC). The cost of energy was included in the utility costs, though. However, the fixed operating and maintenance cost (FOC) also covered the cost of monitoring and other incurred costs. Figure 4.3 shows that scenario 3 had the highest OPEX (36 Million NOK), in contrast to scenario 1, which had the lowest OPEX (25.7 Million NOK). The reason for scenario 3's greater OPEX may be the cost of the utility, which accounts for about 48% of the OPEX. Utility costs accounted for 50% in scenarios 1 and 2, in contrast. It should be noted that the OPEX for each scenario was mostly influenced by utility expenses. Finding a cost-effective utility source is therefore essential for assessing the economic viability of a project.
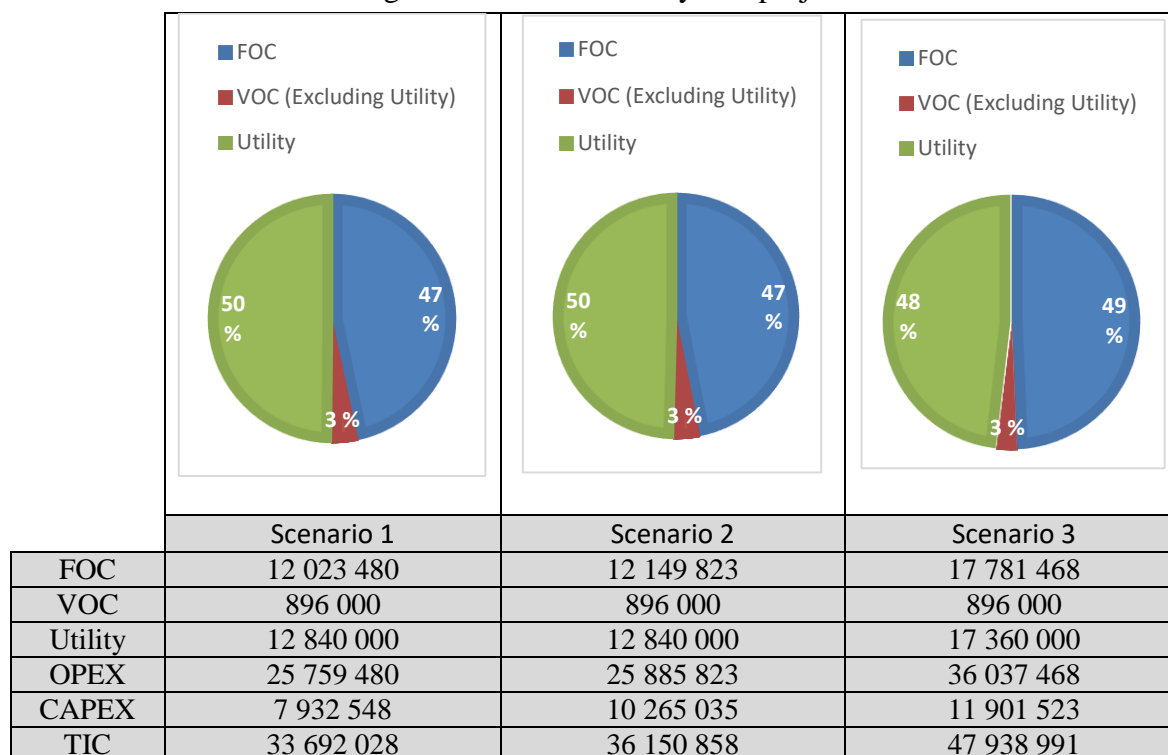

	Scenario 1	Scenario 2	Scenario 3
FOC	12 023 480	12 149 823	17 781 468
VOC	896 000	896 000	896 000
Utility	12 840 000	12 840 000	17 360 000
OPEX	25 759 480	25 885 823	36 037 468
CAPEX	7 932 548	10 265 035	11 901 523
TIC	33 692 028	36 150 858	47 938 991

Figure 4.3 OPEX breakdown for each scenario

4.2.3 Discounted cash flow rate

The fluctuation of the total discounted and undiscounted (present) cash flow over the plant's lifetime is depicted in Fig.4.4 (a-c) for three scenarios. The cumulative non-discounted and discounted cash flow graphs over the plant's lifetime are shown in this picture. Working capital (the additional expenditure required for the plant start-up and initial operation until producing revenues) was set at 5% TIC in this analysis. A lifetime of 20 years and a 15-year linear depreciation were also considered. In addition, the capital investment was split into two equal payments: 50% at the start and 50% at the conclusion of the first year.
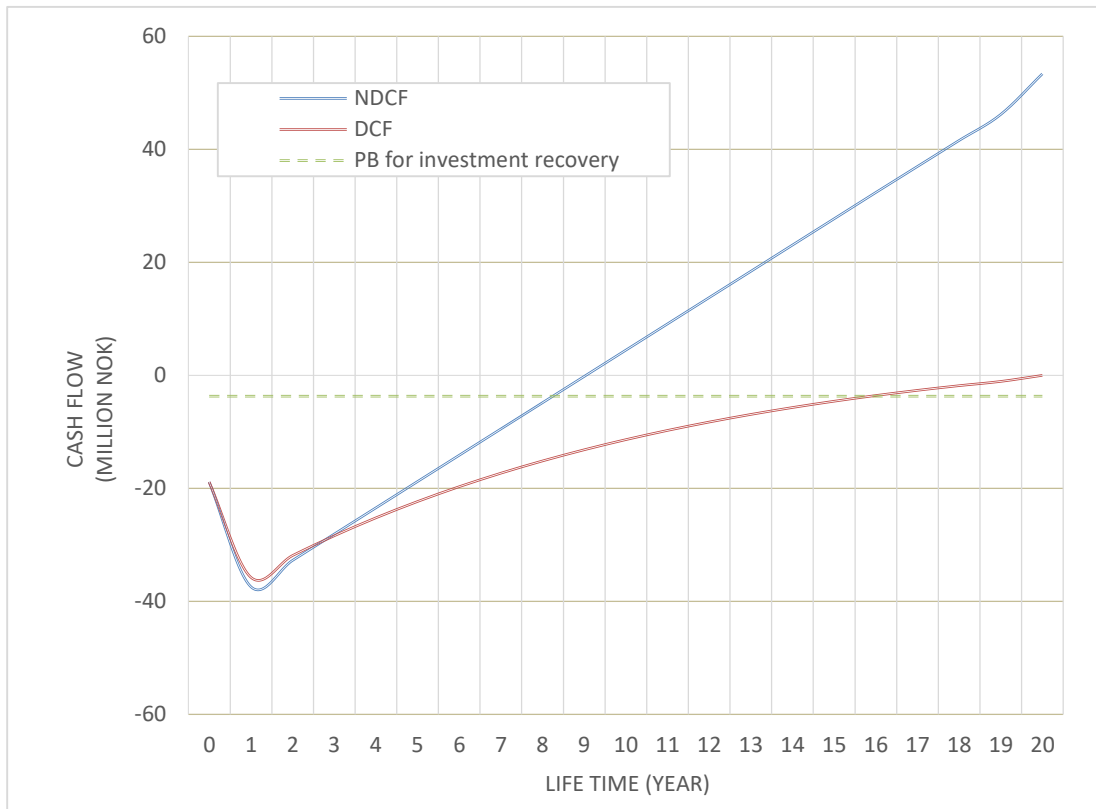
(a)
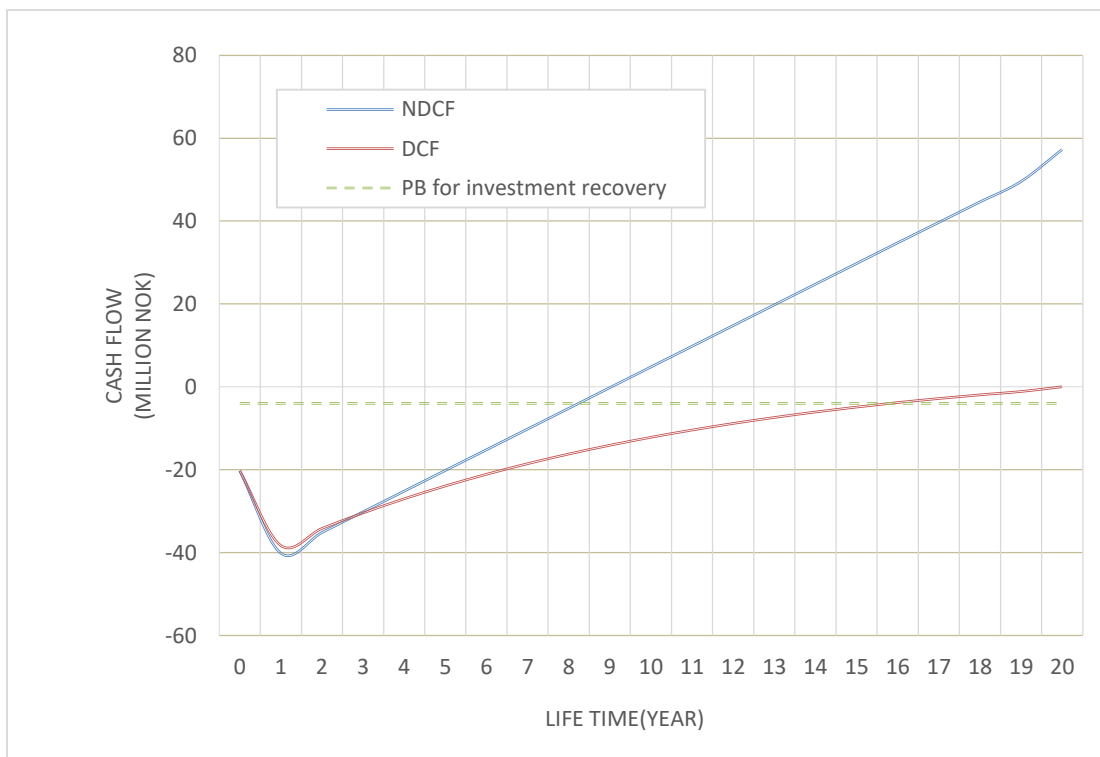
(b)
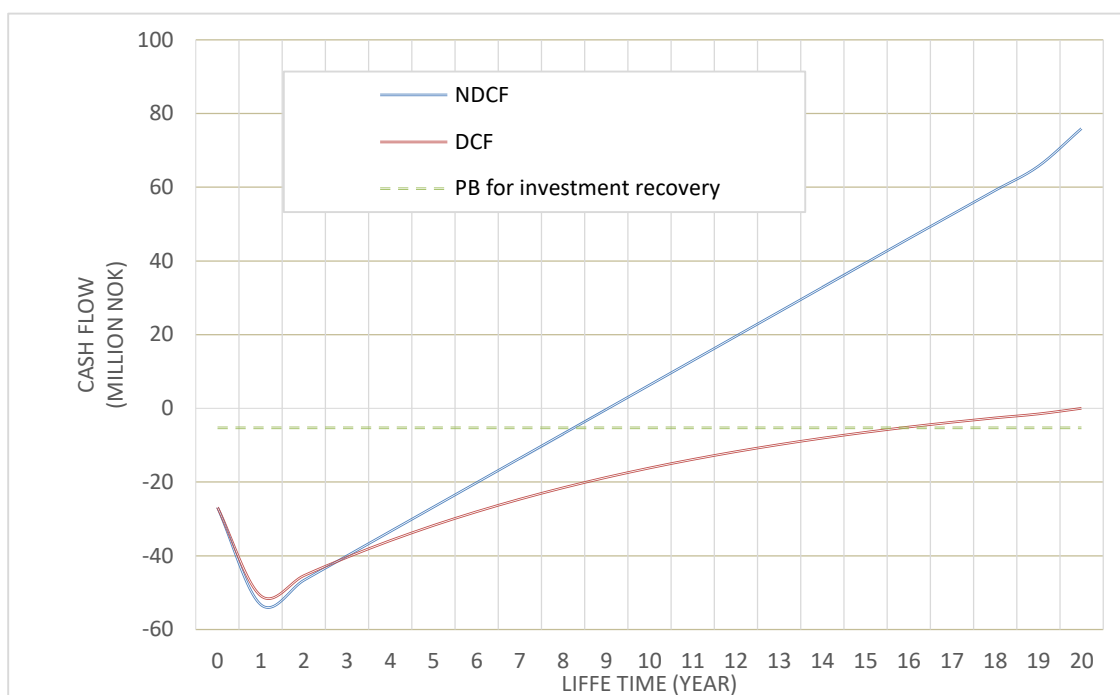
(c)

Figure 4.4 DCFR and NDCFR for Scenario1 (a), Scenario2 (b) and, Scenario3 (c)

Table 4.1. lists the important profitability metrics that were identified.

Table 4.1. Comparison of the main profitability metrics.

Profitability indicators	Scenario 1	Scenario 2	Scenario 3
MSP of Biomethane (NOK/Kg)	237,75	164,64	120,43
Undiscounted NPV (Million NOK)	53,36	57,23	75,91
Discounted NPV (Million NOK)	1,07	1,15	1,52
PBP (year)	8,20	8,50	8,60
DBEP (year)	16,00	16,50	16,50

When compared to scenarios 2 and 1, scenario 3 had the lowest predicted MSP for biomethane (120.43 per kg). In the same way, scenario 3, favored the undiscounted NPV with 75.9 million NOK. While scenario 2 had a NPV (57.2 million NOK), scenario 1 had a smallest NPV of 53.4 million NOK.

4.2.4 Sensitivity analysis

 The effects of the examined factors on the MSP of biomethane for the three scenarios are showed in figure 4.5(a-c). This figure shows the influence of the investigated parameters on the MSP of biomethane for the three scenarios. The cost of labor and utilities had the most effects on the MSP of biomethane. Due to a 30% drop in the price of utility, the MSP of biomethane increased 33.4 NOK/kg in scenario 1 (figure 4.5 - a). The MSP decreased 21.6 NOK/kg , which is comparable to how the cost of labor decreased by 30%. Scenario 2 (figure 4.5 - b) showed a pattern that was identical to scenario 1 with a difference in the cost of the labor and utility. In scenario 2, the MSP of biomethane increased 22.7 NOK/kg and 14.8 NOK/kg, with an increase of 30% in the price of labor and utilities, respectively. The MSP of biomethane in scenario 3 (figure 4.5 - c) significantly changed as a result of changes in utility and labor costs, similar to scenario 2. The MSP of biomethane climbed 16.4 NOK/kg with a 30% increase in utility expenses. When the labor costs increased by 30% , the MSP climbed 11.7 NOK/kg.
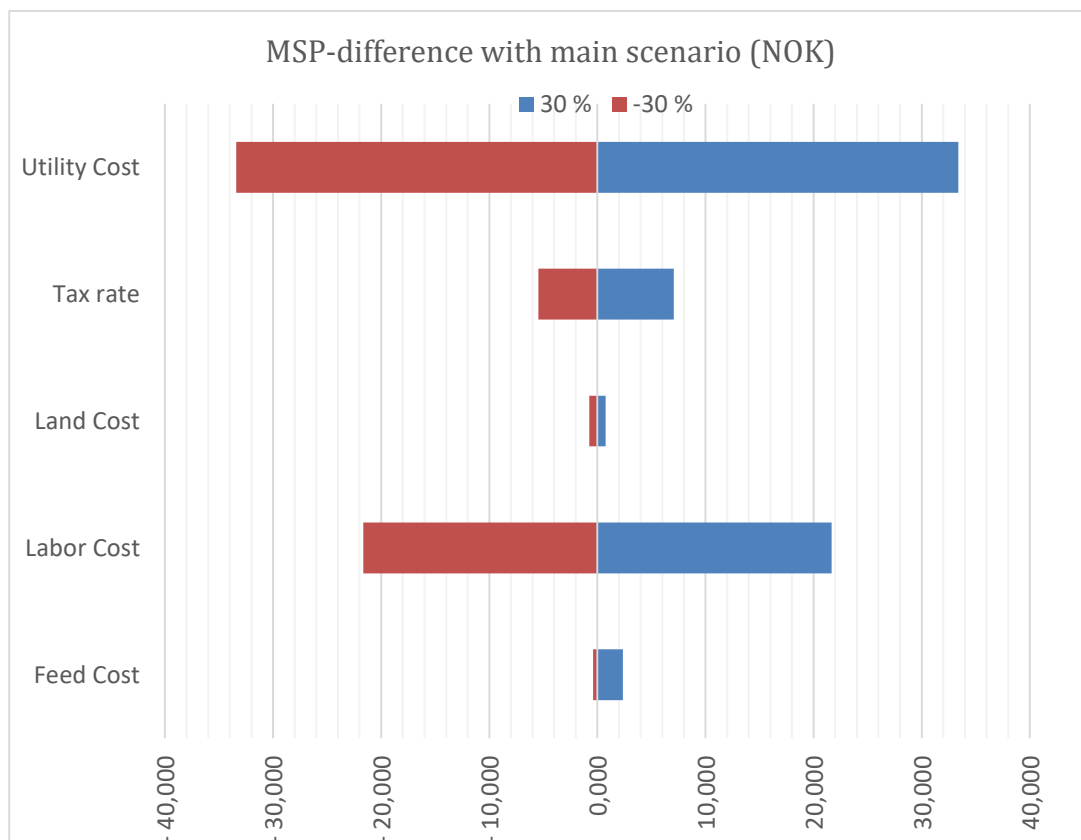
(a)
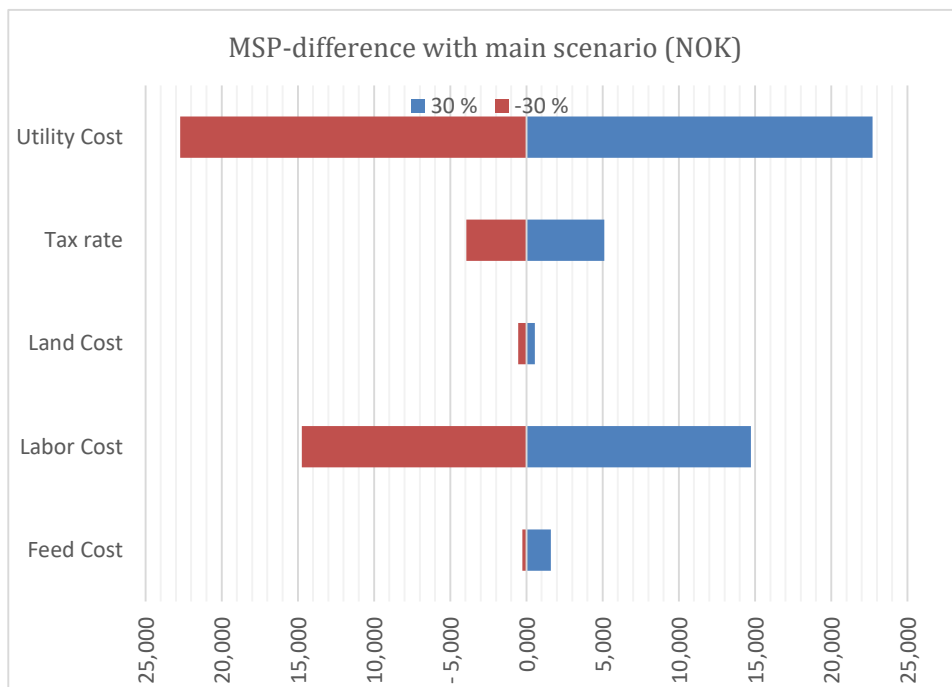
(b)
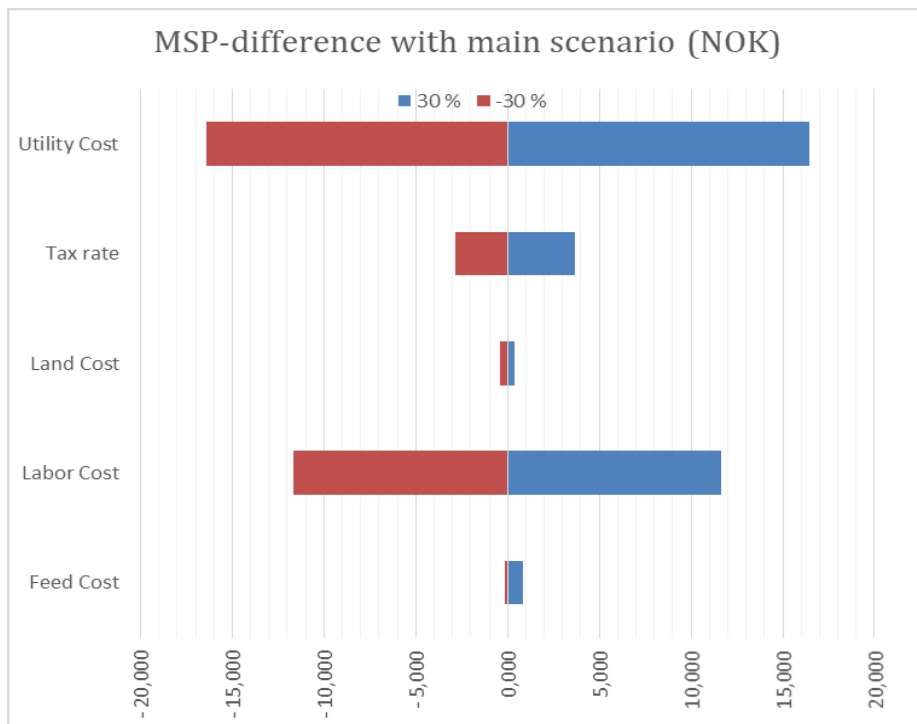
(c)

Figure 4.5 Sensitivity analysis for Scenario1 (a), Scenario2 (b) and, Scenario3 (c)

4.3 Efficiency calculation

In order to reach efficiency of plant for each scenario, a comprehensive energy and mass balance will be required. Energy and mass flow rate of each stream inside the conceptual plant has been determined via Aspen Plus simulation.

4.3.1 Energy and mass balance

Fig. 4.6 (a-c) display the energy and mass balance of each scenario. It should be noted that all the data presented in the following figures are extracted from Aspen plus simulation of each scenario. As previously mentioned, hydrothermal gasification and syngas fermentation were present in every scenario. Similar to this, birch wood was the identical feedstock used in all of the situations. The ratio of the quantity of energy that is usefully output from the products to the energy that is input into the system was used to assess the energy efficiency of each process. The system's total energy input was included in the latter. It should be mentioned that during the energy efficiency calculations, the ratios of each energy output and input were stated as a product of the matching lower heating values (LHV) of each component. Scenario 3 is shows an ideal results in comparison with other scenarios. Scenario 1 and, 2 are also assessed to provide a better comparison. Efficiency of each process is calculated in the next part. Scenario 1 is the simplest scenario due to its simple arrangement. Scenario 2 and 3 compare the effect of adding water and hydrogen latter in the process of biomethane production, respectively. And they prove that adding hydrogen to enrich the fermentation process is more effective than water.
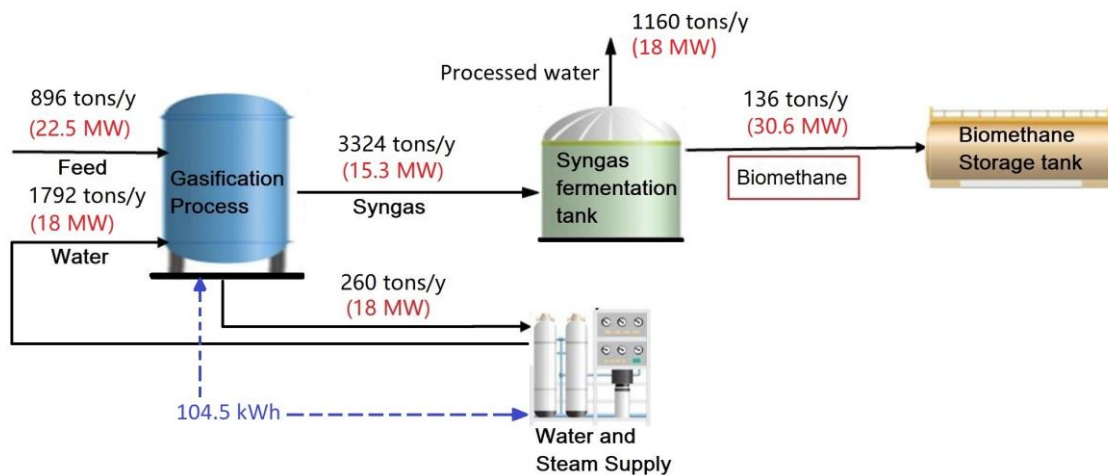
(a)
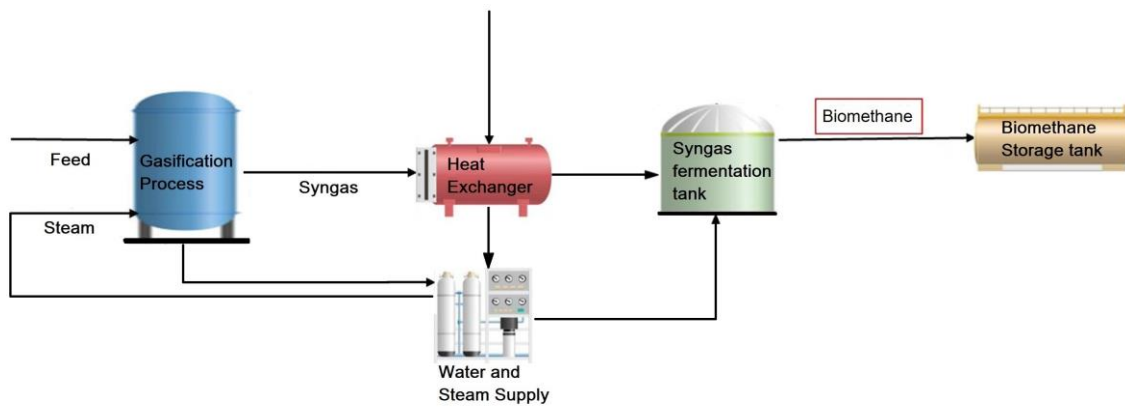
(b)
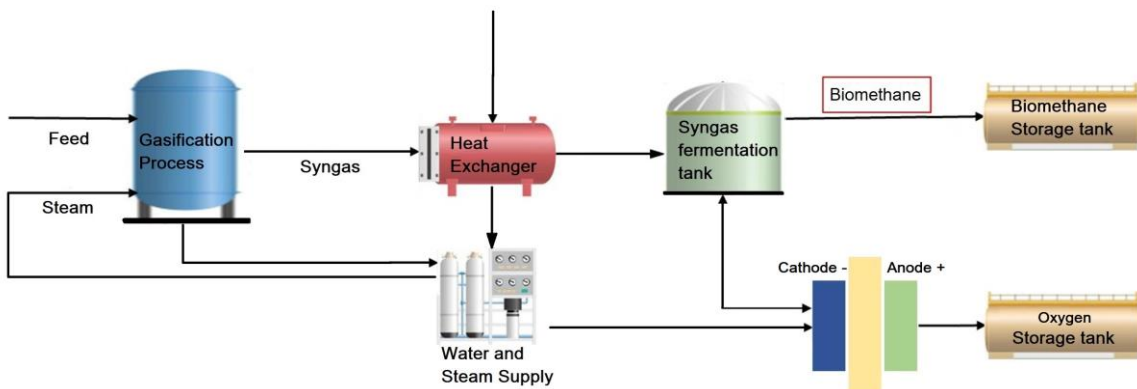
(c)

Figure 4.6 Sensitivity analysis for Scenario1 (a), Scenario2 (b) and, Scenario3 (c)

4.3.2 Plant efficiency

Plant efficiency is a key factor in the assessment of a biogas plant, as it determines the overall effectiveness of the process in converting the feedstock into energy. The efficiency of a biogas plant can be evaluated in terms of energy efficiency.

The energy efficiency of a biogas plant is the amount of energy produced as biogas compared to the energy input required to operate the plant. This metric takes into account the energy used to run the plant, such as electricity for pumps and mixers, and can be affected by factors such as plant design, feedstock type, and operating conditions. A higher energy efficiency indicates a more effective process and can reduce the operating costs of the plant.

The energy efficiency of a biogas plant is the amount of energy produced as biogas compared to the energy input required to operate the plant. This metric takes into account the energy used to run the plant, such as electricity for pumps and mixers, and can be affected by factors such

as plant design, feedstock type, and operating conditions. A higher energy efficiency indicates a more effective process and can reduce the operating costs of the plant.

It should be mentioned that during the energy efficiency calculation, the ratios of each energy input and output were stated as a product of the matching lower heating values (LHV) of each component.

Table 4.2 displays plant efficiency of each scenario according to the correlation stated in the third chapter.

Table 4.2. Plant efficiency of each scenario.

	feed flow rate (kg/hr)	LHV of feed (MJ/kg)	Biomethane flow rate (kg/hr)	LHV of Methane (MJ/kg)	Electricity Consumed (Mwh)	Energy Efficiency
Scenario 1	112	15,023931	17,3	12,55614	0,1045	12,9
Scenario 2	112	15,023931	25,1535	12,55614	0,1045	18,7
Scenario 3	112	15,023931	46,79507	12,55614	0,6695	34,9

5 Discussion

This chapter discusses the results from the simulation, economical assessment and, sensitivity analysis. In order to determine the optimal economic conditions, the impacts of various operating variables on the biomethane production results, are reviewed.

5.1 Process simulation

With a few minor modifications, the proposed model came very near to the experimental findings. All other gas yields generated using the Aspen Plus model marginally outperformed the experimental values, with some few exception. Several factors could be responsible for the minor discrepancy between model predictions and experimental values during the hydrothermal gasification of feed. The heterogeneous nature of birch makes it challenging to model in Aspen plus. The difference between the Aspen Plus simulation findings and the actual feedstock may be due to differences in the feedstock composition. The experimental tests were also carried out in a batch reactor without any agitation [107].

5.1.1 Hydrogen yield

The results showed that higher temperatures improved the generation of hydrogen. According to results presented in figure 4.1 volume fraction of hydrogen produced during the gasification process increases slightly by 2% via enhancement of temperature of the process from 650°C to 850°C. Similarly the same improvement can be observed in experimental data. There is a very small variation between simulation results by Aspen Plus and Experimental data. These differences are 4.5%, 11.1% and, 4.8% for 650°c, 750° and, 850°C, respectively. This trend is also stated in the investigation of Mohammed et al. [15] as discussed in the literature review section.

Hydrogen volume fraction increment by temperature is because that endothermic reforming reaction of hydrocarbons became more effective (as mentioned in chapter 2). On the other hand, variations between experimental and simulation originates from the methodology which is utilized for simulation and assumption which is made during the process of simulation. For hydrogen this range of deviation is completely acceptable and valid.

5.1.2 Carbon dioxide yield

The outcomes demonstrated that carbon dioxide production was reduced at higher temperatures. According to the findings shown in figure 4.1, increasing the process temperature from 650°C to 750°C and 750°C to 850°C causes a 57.4% and 22.35 reduction in the volume fraction of carbon dioxide produced during the gasification process, respectively. Similar reduction trend have also been seen in experimental data. The difference between simulated findings using Aspen Plus and experimental data is incredibly minimal. These variations for 650°C, 750°C, and 850°C are 21%, 26.6%, and 30.3%, respectively. This reduction can also be justified according to Le Chatelier's principle. This behavior is also explained in the study of Lv et al. [38] as stated in the literature review section[38].

5.1.3 Carbon monoxide yield

The findings indicated that greater temperatures enhanced carbon monoxide production. In accordance with the findings shown in figure 4.1, the volume fraction of carbon monoxide produced during the gasification process increases slightly by 5% as a result of raising the process temperature from 750°C to 850°C. The same improvement can also be shown in the experimental data. Between simulated findings from Aspen Plus and experimental data, there is hardly any difference. For 650°C, 750°C, and 850°C, respectively, these disparities are 3.7%, 12.7%, and 24.2%. Same reasons which is mentioned above can be utilized to explain the differences. According to the literature review section, the study by Li et al. [39] explains the pattern as well.

5.1.4 Methane yield

In contrast with other components, there are significant differences between simulation results and experimental data. In accordance with the results depicted in figure 4.1, the volume fraction of methane produced during the gasification process reduced significantly as a result of raising the process temperature from 650°C to 850°C. This reduction behavior is slighter in experimental data. According to the literature review section, this behavior is also explained in the study of Wongsiriamnuay et al. [29].

5.2 Economical evaluation

In this part results of economical and financial concepts are discussed for each of scenarios in different parts such EPC, OPEX, DCFL and, etc. This results have been compared with each other and other availabe literatures.

5.2.1 EPC results

During Scenario 1 the most expensive equipment are pumps and flash separators which allocate 29.4% and, 35.2% of all equipment purchase cost, respectively. These equipment occupy 45.4% and, 27.2% of all equipment purchase cost in scenario 2, respectively .For scenario 3 the most portion belongs to pumps, separators and, electrolyzer which include 39.2%, 23.4% and, 13.7% of all equipment purchase cost, respectively. Electrolyzer in scenario 3 adds extra cost in comparison with scenario 2. EPC of scenario 3 is 50% higher than the EPC of scenario 1 and EPC of scenario 2 is 29.2% higher than the cheapest scenario. In all the scenarios, pumps are the most expensive portion of EPC and Bio-reactor is the smallest one.

5.2.2 Operational and capital expenditure

In contrast to scenario 1, which had the lowest OPEX (25 759 480 NOK), scenario 3 had the greatest OPEX (36 037 468 NOK), as seen in figure 4.3. The cost of utility making up around 48% of the OPEX in scenario 3 may be the cause of its higher OPEX. On the other hand, for scenarios 1 and 2, utility expenses accounted for 50%. It should be highlighted that for all of the scenarios, the OPEX was mostly driven by utility costs. As a result, finding a cost-effective utility source is crucial for determining a project's economic sustainability.

For each scenario, the FOC contribution was likewise substantial. For each case, the FOC contributed between 47% and 49%. In scenarios 1, 2, and 3, FOC made up 47%, 47%, and 49% of the OPEX, respectively.

The CAPEX, which includes fixed capital investment (FCI), working capital, and starting costs, grew as follows: scenario 1 (7 932 548 NOK), scenario 2 (10 265 035 NOK), and scenario 3 (11 901 523 NOK). As a result, finding a cost-effective electricity source is crucial for determining a project's economic sustainability.

5.2.3 Discounted cash flow rate

Fig. 4.4 (a-c) displays the associated cumulative non-discounted and discounted cash flow graphs over the plant's lifespan. In this study, a lifetime of 20 years and a 15-year linear depreciation were assumed, and working capital (the additional investment needed for the plant start-up and initial operation until generating revenues) was set at 5% TIC. Additionally, 50% of the capital investment was paid at the beginning and the remaining 50% at the end of the first year.

In the economic calculation, it was also expected that the plant would not receive any further maintenance due to safety concerns, which resulted in a low salvage value (money from selling the used property at the end of its useful life). Additionally, the factory was built in a year and ran for nineteen years, 330 days a year, with a capacity factor of over 90%. 30% of income was taxed.

Considering Fig. 4.4 (c), the fact that 50% of the TIC and the land are being paid for while the plant is being built results in a negative cash flow in year zero. The most unfavorable cash flow is obtained right at this point because at the end of the first year, 55% of the total investment cost (TIC) (the remaining capital investment plus working capital) are paid, but nothing is created. The investment starts to be recovered at an IRR of 10% when production starts at the beginning of the second year. From this point on and through the last year, there are sales revenues (in this example, from the sale of oxygen and biomethane).

Since depreciation ends after 15 years, the cash flow recovers more slowly from the seventeenth to the final year. The simple non-discounted and discounted payback times, which show the time needed to return the initial investment from the start after discounting the land and working capital, are 8.6 and 16.5 years, respectively. Because the lifespan of the project is longer than its payback period, it can be profitable. Since the non-discounted cash flows change from negative to positive values, the project is generating a return on investment past the non-discounted payback period, which is represented by the sum of all of them (cumulative cash flow) as a positive end value of 76 million NOK.

5.2.4 Sensitivity analysis

The effects of the examined parameters on the MSP of biomethane for the three scenarios are depicted in Figure 4.5 (a-c). The MSP of biomethane was most significantly impacted by the price of utility and labor. The MSP of biomethane increased in scenario 1 from 237.7 NOK/kg to 271.1 NOK/kg as a result of a 30% decrease in the price of utility. Similar to how the cost of labor decreased by 30%, the MSP also changed, going from 237.7 NOK/kg to 216.1 NOK/kg. With a change in the cost of the labor and Utility, scenario 2 demonstrated a pattern that was identical to scenario 1. In scenario 2, with an increase of 30% in the price of utility and labor, respectively, the MSP of biomethane increased from 164.6 NOK/kg to 187.3 NOK/kg and from 164.6 NOK/kg to 179.4 NOK/kg.

Similar to scenario 2, a change in the utility and labor cost led to a substantial shift in the MSP of biomethane for scenario 3. With a 30% increase in utility costs, the MSP of biomethane increased from 120.4 NOK/kg to 136.8 NOK/kg. The MSP increased from 120.4 NOK/kg to 132.1 NOK/kg, while labor expenses increased by 30% at the same time.

The MSP of biomethane was also influenced by other independent factors as tax rate, feedstock cost, and land cost. For the three cases, though, their effects were more muted. It should be underlined that for the three scenarios, the tax rate had the least impact on the MSP of biomethane.

5.2.5 Efficiency of plants

The sequence in which the energy efficiency rose may be observed in each of the figures. Scenarios 1, 2 and, 3 each have three possible outcomes: 12.9%, 18.8%, and 34.9%, respectively. (Table 4.2).

Comparisons were made between the energy efficiency found in this study (19–35%) and that found in the literature on the thermochemical and biochemical conversion of waste materials. For integrated anaerobic digestions, traditional gasification, and biomethanation processes, Michailos et al. [108], showed an efficiency range of 26.5-35.5%.

Energy efficiency for the autothermal reforming and supercritical water reforming of glycerol was reported by Galera and Ortiz [108][109] to be 36% and 35.8%, respectively.

These comparisons states that simulation results has relatively viable and reliable values, especially at third scenario.

5.Discussion

6 Conclusion and recommendation

 Biomass gasification for syngas production and its utilization for biomethane production by coupling anaerobic digestion is one of the best alternatives to valorize the waste material. In this regard, the lab-scale experiments have demonstrated promising results for natural gas-quality biomethane production; however, the economic aspect has often been undermined. Therefore this master's thesis investigated the techno-economical aspect of biomethane production by coupling gasifier and anaerobic digestion. Three different scenarios were evaluated for their techno-economic viability. Scenario 1 is based on biomethane generation through syngas fermentation and hydrothermal gasification. Biomass and steam gasification for hydrogen generation and coupling with syngas fermentation make up Scenario 2. Finally, in scenario 3, an electrolytic unit was added to generate hydrogen in scenario 2 to produce biomethane. Simulation of these scenarios was fulfilled by Aspen Plus software in order to obtain mass and energy balance. By finding the mass and energy balance of each scenario, their efficiencies is defined by simulation results. In addition, the literature review was performed about two processes, including gasification and anaerobic digestion and the optimal operational parameters of each one. By coupling the gasification process with anaerobic digestion, biomethane production from birch waste increased significantly. This coupling has increased the amount of 43 vol.% (individual gasification process) to more than 90 vol.%.

 To determine the designs' economic viability, various profitability indexes were also used to evaluate them. This techno-economic assessment of the economic upscaling of a conceptual process makes it possible for project developers to identify the most important factors affecting the project's economic viability and develop strategies to optimize the process and reduce costs. This can help ensure the long-term sustainability and profitability of the biogas plant. During the economic evaluation, it has been assumed that lifetime of the plants is twenty years, and in the first year entire plant has been supposed to be constructed. Location, currency, and time of evaluation have been considered Norway, NOK and 2023, respectively. The economic viability of the scenarios was assessed using the MSP of biomethane. Following are the declines in MSP for bioethanol: scenarios 1 (237 NOK per liter), 2 (164 NOK per liter), and 3 (120 NOK per liter) are all possible outcomes. The payback period for all the scenarios is almost the same. So. It can be concluded that scenario 3 is the best scenario from economical aspect.

 The analyzed scenarios' efficiencies in terms of energy use range from 13% to 19%. The maximum energy efficiency was found in Scenario 3 at 35%. So, it can be inferred that scenario 3 is the most efficient and profitable scenario among these three scenarios. Scenario 3 has more EPC cost in comparison with other ones. But with its sub-product (oxygen) it compensates the cost faster than the others and reaches profitability period, quickly. The combination of gasification and AD process is an appropriate method to increase the efficiency of each technique. In addition, TEA is one of the most promising methods to evaluate a design and concept before running into a comprehensive biogas plant testing and manufacturing to ensure that your proposal meets the requirements.

 The simulation based on the kinetic model can be studied since this simulation relied on the minimization of the Gibbs free energy model. More parameters, as introduced in chapter two can be investigated. Among those three scenarios, the third one is the most profitable scenario not only for its high biomethane production but also for its high profitability compared with

others. So, it can be chosen as an ideal concept for biomethane production, among others. According to results of EPC section it can be concluded that main portion of each biomethanation plant is devoted to its pumps. As a suggestion to reduce the PBP of third scenario hydrogen can be supported by an external source. This will eliminate electrolyzer which has raised the EPC and subsequently TIC of Scenario 3.

To reduce utility cost which covers a vast portion of OPEX, plant can be located in north of Norway. This will reduce both the Utility and land cost. It can be recommended that by utilizing new methods for automation and artificial intelligence technologies can reduce the labor costs and make the design more profitable and valuable.

During the cash flow rate analysis, it has been supposed that no money is loaned from external source (for example: Bank, etc.). It will make the assessment more realistic if loan amounts and conditions is considered in the TEA.

The construction period has been considered one year that can be last more for larger plants. In addition, feed prices vary depending on a number of variables, including the location of the biomethane plant and the cost of transportation. When maximizing the process and scaling it up, certain parameters should be taken into account.

Furthermore, certain nations' tipping fees, renewable fuel subsidies, and renewable tax credits were not taken into consideration. In any case, the findings of this study provided the framework for the application of technology in the future.

Moreover, effective parameters on MSP of methane were considered. Other factors with minor effect can be significant in other projects that makes them more valuable to be investigated in sensitivity analysis such as working capital and catalyst cost [104].

In future studies, LCA analysis of such scenarios can be valuable from environmental point of view and will add value to the assessment.

The efficiency of a biogas plant can be evaluated not only by energy efficiency, but also by conversion efficiency and overall process efficiency. So, these two recent kind of efficiency can be evaluated for better comparison

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Appendixes

Appendix A
Signed Task Description

University of South-Eastern Norway

Faculty of Technology, Natural Sciences and Maritime Sciences, Campus Porsgrunn

FMH606 Master's Thesis

Title: Techno-economic assessment of biomethane production from syngas derived from biomass gasifier

USN supervisors: Nabin Aryal (main supervisor) and Lars Erik Øi (co-supervisor)

External partner: None

Task background:
Biomethane is a flexible renewable source that can be applied to substitute fossil-based natural gas. Due to the current geopolitical situation, biomethane has recently been proposed as an alternative gas source for Europe. As a result, the number of biogas plant installations in Europe has been increasing for the utilization of waste to produce methane. However, complete anaerobic degradation of organic waste is not possible due to the presence of lignocellulose biomass. Therefore, high lignocellulose-containing material, particularly undigested fibrous material from AD, forest and agriculture biomass, can be utilized in thermal degradation to generate the intermediate gas composition called syngas. The syngas often contains carbon dioxide, carbon monoxide and hydrogen. These gaseous compounds could further utilize to produce natural gas-quality biomethane (>97% methane) in anaerobic digestion.

The two-stage (gasification-AD) process for biomethane has recently received extensive attention and has been researched comprehensively. However, the cost of biomethane production has always been questioned while upscaling the technology. Therefore, this Master thesis will perform the techno-economic evaluation of a two-stage process for biomethane production. In this proposed master's thesis, the student will develop a model based on the available data from the literature for economic assessment. The production cost estimation per cubic meter of methane will be expected using a biomass gasifier coupled with anaerobic digestion.

Task description:
The primary aim of this project work is: -
1. To perform the techno-economic evaluation of a two-stage (gasification-AD) biomethane production process
2. To propose the economic methane production rate scenario for upscaling the technology

The following experimental design will be adopted to achieve the proposed objectives:
The literature review will be done based on the published result from biomass and gasification coupled processes. The thesis should define the i) Designing of a process framework for biomethane production ii) Establishing the mass and energy balances based on the designed framework, iii) total cost estimation based on the capital expenditure (CAPEX) and operating expense (OPEX) and iv) Sensitivity analysis. The student is supposed to estimate the scenario of a profitable biomethane production rate based on the current lab-scale production rate.

Student category: EET or PT

Is the task suitable for online students (not present at the campus)? Yes

Practical arrangements:
The available simulation tool and model will be provided that is available at USN, Porsgrunn campus.

Supervision:
As a general rule, the student is entitled to 15-20 hours of supervision. This includes the necessary time for the supervisor to prepare for supervision meetings (reading material to be discussed, etc.).

Signatures:

Supervisor (date and signature): *nabin aryal*   15.01.2023

Student (write clearly in all capitalized letters):   MILAD SAHRARO

Student (date and signature): *Milad Sahraro* 15.01.2023

Appendix B

Table 2.1 Temperature effect on gasification process

Continued

NO.	Gas yield range						Gas composition				
	TG	**H2**	**CO**	**CH4**	**CO2**	**Unit**	**H2**	**CO**	**CH4**	**CO2**	**Unit**
1	0,5-1,55	0,175-0,7	0,125-0,42	0,05-0,12	0,125-0,3	(m3/kg biomass)	na	na	na	na	na
2	62,68-91,7	-	-	-	-	wt%	10,27-38,2	21,87-36,36	5,84-14,72	10,0-65	vol %
3	1,43-2,53	-	-	-	-	(Nm3/kg biomass)	21-39	35-43	6,0-10	18-20	vol %
4	1,79-2,48	0,861-1,481	-	-	-	(m3/kg)	48-60	15-25	5,0-5,0	20-25	vol %
5	1,9-2,0	-	-	-	-	(Nm3/kg biomass)	6,6-8,16	23,5-30,6	4,0-5,0	59-63	vol %

Continued

NO.	CCE (%)	LHV (MJ/m3)	OC					RN
			BT1 (°C)	description 1	BT2 (°C)	description 2		
1	na	na	800	Highest H2 and gas yield, and low char and tars	-			[37]
2	na	7,5-15,55	1000	Highest LHV and gas yield, and low char and tars	-			[15]
3	78,17-92,59	7,362-8,56	900	Highest CCE and gas yield, and low char and tars	800	Highest LHV		[38]
4	na	9,13-11,26	900	Highest H2 and gas yield	750	Highest LHV		[39]
5	63,6-67,4	1,6-1,9	400	Highest H2	500	Highest LHV and CCE		[29]

Continued

NO.	System configuration and operation parameters														
	GT	H (mm)	ID (mm)	F	FR (kg/h)	FS (mm)	BM	GA	SB	ER	BT (° C)				
1	CFB	NA	60	almond shells	0,06	0,287-1,09	fine Alumina	steam	0,8	na	600	650	700	750	800
2	FLB	NA	40	EFB (<10 wt% moisture content)	0,6	0,3-1	inert sand	air	na	0,15-0,35	700	800	900	1000	-
3	FLB	1400	40	pine sawdust	0,445	0,3-0,45	Silica sand	air-steam	2,7	0,22	700	750	800	850	900
4	FIB	1200	88	palm oil wastes	0,3-1	0,15-2	tri-metallic catalyst	steam	1,33	na	750	800	850	850	900
5	FLB	2000	50	bamboo	0,6	0,1-0,25	Silica sand	air	na	0,4	400	500	600	-	-

Table 2.2 Main reactions on gasification process [12]

Reaction	Heat of reactiona	Reaction name	Reaction number
Heterogeneous reactions:			
$C+0.5O_2=CO$	−111 MJ/kmol	Char partial combustion	R1
$C+CO_2 \leftrightarrow 2CO$	+172 MJ/kmol	Boudouard	R2
$C+H_2O \leftrightarrow CO+H_2$	+131 MJ/kmol	Water−gas	R3
$C+2H_2 \leftrightarrow CH_4$	−75 MJ/kmol	Methanation	R4
Homogenous reactions:			
$CO+0.5O_2=CO_2$	−283 MJ/kmol	CO partial combustion	R5
$H_2+0.5O_2=H_2O$	−242 MJ/kmol	H2 partial combustion	R6
$CO + H_2O \leftrightarrow CO_2+H_2$	−41 MJ/kmol	CO shift	R7
$CH_4 + H_2O \leftrightarrow CO+3H_2$	+206 MJ/kmol	Steam-methane reforming	R8
Hydrogen sulfide (H2S) and ammonia (NH3) formation reactions:			
$H_2+S=H_2S$	nr	H2S formation	R9
$0.5N_2+1.5H_2 \leftrightarrow NH_3$	nr	NH3 formation	R10

Table 2.3 Effect of particle size on the gasification process

NO.	System configuration and operation parameters										
	GT	H (mm)	ID (mm)	F	FR (kg/h)	FS (mm)	BM	GA	SB	ER	BT (ºC)
1	FLB	NA	40	EFB (<10 wt% moisture content)	0,6	<0,3 & 0,3-0,5 & 0,5-1,0	inert sand	air	na	0,15-0,34	850
2	FIB	1200	88	palm oil wastes	0,3-1	0,3-1	tri-metallic catalyst	steam	1,33	na	850
3	FLB	1400	40	pine sawdust	0,512	0,2-0,9	Silica sand	air-steam	1,56	na	800
4	FIB	600	219	pine sawdust	0,3	<0,075 & 0,075-1,2	calcined dolomite	steam	na	1,2	900
5	FIB	1000	50	CDCB	NA	<0,15 & 0,15-3,0	NA	steam	na	na	850

Continued

NO.	Gas yield range			Gas composition					CCE (%)
	TG	H2	Unit	H2	CO	CH4	CO2	Unit	
1	72,74-74,79	na	wt%	21,57-33,93	35-42,5	15-17,5	7,5-30	na	na
2	2,16-2,41	1,183-1,4	(m^3/kg)	55-58	14-18	3,0-5,0	20-23	vol $^\%$	na
3	1,53-2,57	na	(Nm^3/kg biomass)	30-32	16-20	6,0-7,0	16-20	vol $^\%$	77,62-95,10
4	1,38-1,62	0,55-0,8	(m3/kg biomass)	40-51,2	15-22,4	2,0-5,0	12,0-40	vol $^\%$	80-99,87
5	1,72-1,84	1,72-1,84	(Nm^3/kg biomass)	47-49	14-15	2	30-35	vol $^\%$	na

Continued

Continued

NO.	LHV (MJ/m3)	OC					RN
		FS1 (mm)	description 1	FS2 (mm)	description 2	
1	11,8-15,26	<0,3	Highest gas yield, and low char and heavy tars	0,3-0,5	Highest LHV	[15]
2	8,99-10,28	<0,15	Highest H_2 and gas yield	2,0-5,0	Highest LHV	[39]
3	7,0-8,7	0,2-0,3	Highest CCE, LHV and gas yield	-	-	[38]
4	na	<0,075	Highest H_2, CCE and gas yield, and low char and tars	-	-	[27]
5	na	0,45-0,9	Highest gas yield	-	-	[42]

Table 2.4 Effect of gas agent on the gasification process

NO.	System configuration and operation parameters								
	GT	H (mm)	ID (mm)	F	FR (kg/h)	FS (mm)	GA (FFA)	GA (FSA)	BT (° C)
1	FIB	350	60	pine sawdust	NA	NA	Air	Oxygen-Steam	-
2	FLB	NA	50	bamboo	0,6	0,1-0,25	Air	Air-Steam	400,500,600
3	BFB	NA	200	NA	NA	NA	Air	Pure steam	-
4	FIB	400	NA	palm oil wastes	0,3	0,15-2	Steam	na	800
5	FLB	1100	63,9	alfa cellullus	NA	<0,35	Air-Steam	na	800
6	FLB	NA	40	EFB	0,6	0,3-0,5	Air	na	850
7	FIB	500	12,5	Olive Kernel	NA	NA	Air	na	950

Continued

Continued

NO.	Gas yield range					Gas composition				
	TG (FFA)	H2 (FFA)	TG (FSA)	H2 (FSA)	Unit	H2	CO	CH4	CO2	Unit
1	0,82-0,94	0,24-0,33	1,24-1,62	0,36-0,49	(Nm3/kg) (m3/kg)	21,57-33,93	35-42,5	15-17,5	7,5-30	na
2	1,9-2,0	NA	1,4-2,4	NA	(Nm3/kg)	55-58	14-18	3,0-5,0	20-23	vol %
3	1,4-2,4	NA	0,8-1,1	NA	(Nm3/kg biomass)	30-32	16-20	6,0-7,0	16-20	vol %
4	1,2-2,48	0,558-1,481	NA	NA	(m3/kg)	40-51,2	15-22,4	2,0-5,0	12,0-40	vol %
5	0,78-1,02	NA	NA	NA	(m3/kg)					
6	70,75-86,46	NA	NA	NA	wt%					
7	0,6-0,8	NA	NA	NA	wt%	47-49	14-15	2	30-35	vol %

Continued

NO.	CCE (%) (FFA)	CCE (%) (FSA)	LHV (MJ/m3)	OC	RN
1	na	na	na	Oxygen steam	[43]
2	63,6-67,4	87,3-95,5	na	air steam ER:0,4 SB 1:1	[29]
3	na	na	na	Air	[12]
4	na	na	8,73-11,98	SB: 1,33 for maximum H2 SB:0 for (Highest LHV)	[39]
5	na	na	6,55-7,61	SB: 1 for maximum H2 SB:0 for (Highest LHV)	[44]
6	na	na	12,35-15,38	ER:0,35 biogas Max ER:0,25 H2 Max ER:0,15 LHV Max	[15]
7	na	na	8,8-10,4	ER:0,21 biogas Max H2 Max LHV Max	[40]

Table 2.5 Bed material effect on the gasification process

NO.	GT	H (mm)	ID (mm)	F	FR (kg/h)	FS (mm)	GA	BT (ºC)	BM
									System configuration and operation parameters
1	BTR	NA	-	pine bark	NA	NA	Steam	600	CaO reagent as CO2 Sorbent CaO/B:1
2	BTA	NA	-	Japanise oak	NA	NA	Steam	650	Ca(OH)2 powder as CO2 Sorbent
3	FIB	500	10,5	glycerol	NA	0,2-0,35	Steam	800	Silicon Carbide and Ni/Al2O3catalyst
4	FLB	2000	50	Bamboo	0,6	0,1-0,25	air	400-600	Silica sand and calcined dolomiteas catalyst
5	FIB	1200	88	palm oil waste	0,3-1	0,15-2	Steam	800	no catalyst, calcineddolomite,nano-NiLaFe/γ-Al2O3

Continued

ON.	TG (WBM)	H2 (WBM)	TG (WOB)	H2 (WOB)	Unit	H2(WBM)	CO(WBM)	CH4(WBM)	CO2(WBM)	H2(WOB)	CO(WOB)	CH4(WOB)	CO2(WOB)	Unit
	Gas yield					Gas composition								
1	1,42	-	0,87	-	(m3/kg)	64,5	5,9	2,8	26,8	60	9,1	3,2	27,7	vol %
2	NA	0,5-0,8	NA	NA	(m3/kg)	na	na	na	na	na	na	na	na	vol %
3	0,91-1,3	NA	NA	NA	(m3/kg)	55,4-68,3	20,2-36,9	2,4-5,9	1,9-7,7	na	na	na	na	mol %
4	1,9-2,1	NA	NA	NA	(Nm3/kg)	3,2-9,1	21,4-31,7	na	na	na	na	na	na	mol %
5	1,21-2,11	0,442-1,131	NA	NA	(m3/kg)	36,5-53,6	12,7-25,8	4,4-10,2	20,9-26,6	na	na	na	na	vol %

Continued

Continued

NO.	CCE (%) (WOB)	CCE (%) (WBM)	LHV (MJ/Nm)	OC		RN
1	30,3	55,6	na	Using CaO		[48]
2	na	na	na	[Ca]/[C]: 2 (highest H2 yield)		[49]
3	na	na	na	Catalyst loading: 0.8 wt% (highest H2 content and lowest char content)		[53]
4	59,5-80,1	na	na	C/B: 1.5:1 (maximum H2 content, Highest LHV, and highest CCE)		[29]
5	na	na	10,2-12,72	BM: newly developed tri-metallic catalyst (maximum gas yield, maximum H2 yield, and optimum gas composition)	BM: calcined dolomite (highest LHV)	[39]

Table 2.6 Effects of different factors on anaerobic digestion

Environmental factors	Amounts	Target substrates	Performance	RN
Temperature	288 K	Rice	26-day digestion periods	[63]
	308 K	Pig manure	30-day digestion periods	[64]
	308 K	Wheat straws	20-day digestion periods	[64]
pH	4.5 – 5.5	Corn straws	Enhance acid production	[70]
	<4 or >6	Cotton stalks	No obvious effects	[71]
	<6 or >7.8	Left vegetables	Inhibit biogas production	[77]
	7	Pig manure	Greatly enhanced biogas production	[77]
Nitrogen	$0.00357 – 0.0357$ mol L^{-1}	Left fish and vegetables	No obvious effects	[77]
	7.14×10^{-3} mol L^{-1}	Human manure	Slightly inhibited digestion	[79]
	0.028 mol L^{-1}	Cow manure	Greatly inhibited digestion	[78]
OLR	$5 – 10$ kgCOD m^{-2}.day^{-1}	Sugar cane stalks	No obvious effects	[83]
	$13.63 – 19.27$ kgCOD m^{-2}.day^{-1}	Cotton stalks	Enhanced biogas production	[82]
	> 30 kgCOD m^{-2}.day^{-1}	Cow manure	Inhibitory effects	[82]

Continued

Environmental factors	Amounts	Target substrates	Performance	RN
LCFAs	0.8 – 0.9 kg m^{-3}	Noodles and meats	Negative influences	[89]
	>1 kg m^{-3}	Wheat straws	Greatly inhibited digestion	[88]
	>1.5 kg m^{-3}	Human manure	AD system fails	[87]
Light metallic ions	0.1 – 1 kg m^{-3} IT	NA	NA	[93]
	1–2 kg m^{-3} IT	Left eggs and meats, rice	No obvious effects	[91]
	>2.5 kg m^{-3} IT	Pig manure	Inhibit acid and biogas formation	[92]
Heavy metallic ions	0.0005 – 0.003 kg m^{-3} IT	Chicken manure	Slightly enhanced biogas formation	[95]
	0.004 – 0.1 kg m^{-3} IT	Sugar cane stalks	Toxic to bacteria and methanogens	[92]
	0.0156 kg m^{-3} IT	Left noodles and rice	AD system fails	[95][88]

Appendix C

Table 3.6 Chemical Engineering Plant Cost Index for various years

No.	Name	Price (million USD)	Ref Year	CEPCI	CEPCI 2023
1	Bioreactor for syngas fermentation	0,91	2016	541,7	765,8
2	Electrolysis unit	1,45	2017	567,5	765,8
3	Flash separators	0,11	2014	576,1	765,8
4	Heat exchangers	0,31	2014	576,1	765,8
5	Gasification reactor	0,59	2019	620,2	765,8
6	Pumps	0,12	2014	576,1	765,8

Table 3.7 Data utilized for EPC calculation [101]

No.	Name	Ref Capacity	Current Capacity	Unit	Scaling factor
1	Bioreactor for syngas fermentation	7695	0,495	Volume (m3)	0,6
2	Electrolysis unit	1	0	Installed capacity (MW of electricity)	0,85
3	Flash separators	10	2	Length (m)	0,78
4	Heat exchangers	1000	100	Area (m2)	1
5	Gasification reactor	56	0,95	Feed input (tons/year)	0,72
6	Pumps	10	1	Power (kW)	0,36

Table 3.11 Discounted cash flow rate calculation sheet [100]