

UNIVERSITY OF SOUTHERN DENMARK

BACHELOR THESIS

Implementation of Fischer-Tropsch Jet Fuel Production in the Danish Energy System

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Declaration of Authorship

I, Sebastian Frejo Rasmussen, declare that this thesis titled, *Implementation of Fischer-Tropsch Jet Fuel Production in the Danish Energy System* and the work presented in it are my own. I confirm that:

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Abstract

This bachelor thesis focuses on the prospects of the implementation of Fischer-Tropsch jet fuel production in Denmark. It examines this from a technical and economic viewpoint under the assumption that domestic production must satisfy the entire Danish jet fuel demand. Simulations of a model, made in Aspen Plus, provide the performance of the Fischer-Tropsch gas-to-liquids conversion, which form the basis of the analysis. Based on the simulations, the study investigates three scenarios with different advantages and disadvantages regarding their implementation into the Danish system. The three scenarios are; one decentralised scenario where the plant feed is biogas, and two scenarios with a centralised production both with a plant feed of biomethane but one of them with co-feed of CO₂. The results of the analysis show that the scenario with the best economic prospects is the scenario with decentral production plants. However, technical and structural complications make it the more uncertain solution. All of the evaluated scenarios rely on subsidies, and all of them will consume close to the total Danish potential of biogas or biomethane. Satisfying the entire Danish jet fuel demand from Fischer-Tropsch jet fuel, therefore, requires careful and strategic energy planning.

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List of Abbreviations:

ATR: Autothermal Reforming

bbl: barrel

bpd: barrels per day

CAPEX: Capital Expenses

DAC: Direct Air Capture

DEA: Danish Energy Agency

DMR: Dry Methane Reforming

FVF: Fjernvarme Fyn

GTL: Gas-To-Liquids

HSE: Health, Safety, and Environment

HTFT: High Temperature Fischer-Tropsch

LCOE: Levelised Cost of Energy

LHV: Lower Heating Value

LTFT: Low Temperature Fischer-Tropsch

NPV: Net Present Value

OPEX: Operational Expenses

POX: Partial Oxidation

PSA: Pressure Swing Adsorption

R&D: Research and Development

SDR: Steam and Dry Reforming

SMR: Steam Methane Reforming

SNG: Synthetic Natural Gas

SOEC: Solid Oxide Electrolyser Cell

Syncrude: Synthetic crude oil

Syngas: synthesis gas

WTP: Willingness-To-Pay

1 Introduction

In line with the Paris agreement, Denmark should be a net zero-emission society by 2050. The goal of the Danish Government is to make Denmark independent of fossil fuels by this time [1]. The Danish Government made a broad energy agreement (as of June 29th 2018) with all other parties in the Danish Parliament where energy political goals are defined [2]. With this agreement, the Danish politicians want to have accomplished a complete phaseout of coal in the energy system by 2030. For the electricity sector, 100% should be covered by renewable energy, for the district heating sector it should be 90%, but for the transport sector no clear goals have been defined [2]. Likewise, the EU lacks clearly defined targets for the transport sector after that for 2020 saying that 10% of the energy in the sector should be renewable [1]. The lack of ambitious goals within a time horizon of a few years indicates that the transport sector is facing greater problems in the green transition than e.g. the household sector. Regarding personal transport, the Danish Energy Agency (DEA) expects that electric cars will become competitive to traditional cars by 2030, and also electrification of the railway seems straightforward. However, it is not as easy when it comes to road freight, sea freight, and aerial transport. All of these require fuel with a high energy density similar to the oil products used today. Alternatives include biodiesel, synthetic natural gas (SNG), and biokerosene [3]. According to the DEA, the most inflexible of these categories is the aerial transport with biokerosene as the only alternative to conventional kerosene, although, research and development (R&D) related to electrified short distance aeroplanes is also made [4]. The DEA proposes two options when the aerial transport should be based on renewable sources: (1) import of biokerosene and (2) production of biokerosene in Denmark. The second option can have the benefit of a large amount of recoverable heat that can be used in the Danish district heating system. A benefit that is not associated with the first option [3]. On top of that, *SDU Life Cycle Engineering* under the Department of Chemical Engineering, Biotechnology, and Environmental Engineering at SDU is working together with several companies with the aim of constructing a demonstration plant for bio jet fuel production. This plant should use the well-known Fischer-Tropsch synthesis to make jet fuel from methane. For it to be bio jet fuel, the methane can originate from biogas. In line with this, it is interesting to investigate what happens if this is scaled to cover the total demand for jet fuel in Denmark and how this can be implemented in the Danish energy system.

1.1 Aim of the study

This bachelor project focuses on the production of Fischer-Tropsch jet fuel for the entire Danish aviation sector in relation to the Danish energy system. More specifically, to answer questions

about the district heating potential from the overall process, the biogas or biomethane availability, the cost of the bio jet fuel, and the economic aspects associated with this pathway. Furthermore, it will be investigated whether it is more beneficial to have centralised or decentralised production, and if the technically best solution is also the economically best solution. With these questions, the implementation of Fischer-Tropsch jet fuel production in the Danish energy system is examined.

1.2 How the study will be conducted

Aspen Plus will be used to model GTL conversion with methane reforming and a Fischer-Tropsch synthesis to be able to determine the performance of the conversion technology in relation to product efficiency as well as district heating efficiency. On this basis, three scenarios are analysed that differ on the feed to the GTL plants. They are biogas, biomethane, and biomethane with CO₂ co-feed. These three scenarios will be evaluated in the Danish energy system in a time horizon from today until 2050.

1.3 Scope and delimitation

The model in Aspen Plus will be made with simplifications in regards to the involved products and by-products. The analysis will be made in a broad system perspective. This means that the results cannot be used as a direct guide to exactly how and where GTL plants should be placed in Denmark. The study is a theoretical work based on previous findings but will add to the subject how the production of Fischer-Tropsch jet fuel fits into an energy system like the Danish when it is to satisfy the total demand.

1.4 Reading guide

Background data, relevant definitions, technologies, and other prerequisites in relation to the analysed part of the energy system are described in Section 2. Section 3 describes the methods and simplifications in the project in relation to the background data. Section 4 shows the modelling in Aspen plus, and the main simulation results are given in Section 5. The three analysed scenarios are described in Section 6 followed by the analysis and discussion in Section 7. In Section 8 the conclusion of the bachelor project is given.

2 Background data and information

In this section, the analysed case is put in context in regards to the technology and the Danish energy system, and data used in the study will be presented.

2.1 Fischer-Tropsch GTL technologies

One of the ways to produce biokerosene is through methane reforming and Fischer-Tropsch synthesis. This pathway is looked further upon in this study. The method is basically to convert methane into syngas by methane reforming in a reformer which is then synthesised into long-chained liquid hydrocarbons in a Fischer-Tropsch reactor. Methane is present in biogas and in upgraded biogas, i.e. biomethane.

For the conversion technologies, defining the reformer for the syngas production and the type of Fischer-Tropsch synthesis reactor is interesting. This is because they are the main components in the GTL conversion [5, 6]. Typical process flow sheets for GTL plants make the foundation of the system setup in the current study. These process flow sheets can be found in several articles and books on the topic [5, 7, 8, 6].

2.2 Definitions

A brief definition of the efficiencies used to compare the different setups follows.

Carbon efficiency: The carbon efficiency is the ratio between C-atoms in the liquid products and the feed.

Thermal efficiency: The efficiency evaluated based on the lower heating values (LHV) of the products and feed.

Product efficiency: The product efficiency is the thermal efficiency evaluated as the ratio between the liquid products and the feed with the utilisation of tail gases.

District heating efficiency: The thermal efficiency evaluated as the ratio between the recoverable heat and the energy in the feed with the utilisation of tail gases. The district heating efficiency is the difference between the total efficiency and the product efficiency.

Total efficiency: The total efficiency is the sum of the product efficiency and the district heating efficiency.

Some of the most important ratios in relation to the subject discussed in this report are.

H₂O:CH₄: The water to methane molar ratio that is fed to the reformer.

CH₄:CO₂: The methane to carbon dioxide molar ratio that is fed to the reformer.

H₂:CO: The hydrogen to carbon monoxide that is present in the syngas after the reformer and before the Fischer-Tropsch synthesis.

These ratios are important for the GTL conversion as they are determinants of the performance and the output of the process.

2.2.1 The reformer

The catalytic process of methane reforming is the most common way to produce syngas [9]. The main technologies for this are

- Autothermal Reforming (ATR).
- Partial Oxidation (POX).
- Steam Methane Reforming (SMR).
- Dry Methane Reforming (DMR).
- Combined Steam and Dry Reforming (SDR).

Common for the first two reformer types, ATR and POX, is that both of them require pure oxygen in the process which is expensive to extract from air [10, App. G]. The advantage of these two is that the obtained H₂:CO ratio lies within the range of typical usage ratios for Fischer-Tropsch synthesis of 1.6-2.15 [5, ch. 3]. The SMR, on the other hand, yields an H₂:CO ratio well above 3, which is way too high compared to the usage ratio in the Fischer-Tropsch synthesis no matter the design, although, this technology only uses steam. This can be adjusted in the conditioning process of the syngas but results in a waste of energy. DMR uses neither oxygen nor steam but CO₂ for the reforming and gives an H₂:CO ratio of 1 which is too low. The combination of these two technologies is the SDR which gives a variable H₂:CO ratio so the Fischer-Tropsch requirements can be met. It requires CO₂ though, which must be captured before it can be co-fed unless it is already present in the feed [5, 7].

Regarding the choice of the reformer, it is decided only to look at the SMR and SDR reformers as they are regarded as the most appropriate options to avoid the extraction of pure oxygen from the air. The operating conditions for the SMR are based on *A. de Klerk* [5, ch. 3] and for the SDR, they are based on the description from *I. S. Ermolaev* [7].

Syngas can also be produced from other sources than methane. Co-electrolysis is one way where the electrolysis of water to H_2 can be accompanied by the electrolysis of CO_2 to CO [11]. This is interesting because a possible methane shortage can be avoided. One technology for this is the solid oxide electrolyser cell (SOEC). The SOEC technology is, however, rather immature and the DEA does not have prices for it in 2015, and in 2020 the prices are around four times higher than expected in 2030 [12, Technology: SOEC]. This technology is, therefore, not relevant today, but with future developments, it might become interesting in the energy system later.

2.2.2 The Fischer-Tropsch synthesis

There are four main Fischer-Tropsch synthesis configurations. They are high- and low-temperature technologies (HTFT and LTFT). Typically, they use either a cobalt catalyst or an iron catalyst, although other catalysts exist. A term used to describe the probable products in the syncrude is the *probability of chain growth*, α . The greater this is, the greater is the probability of getting heavier hydrocarbons. α is greater for LTFT than for HTFT, which means that the main products from LTFT synthesis typically are wax, diesel, jet fuel, and naphtha, while they are petrol and petrochemicals for HTFT synthesis. These overall synthesis technologies can be fine-tuned even further so the amount of the desired products is increased [13]. It is, in general, better to use LTFT than HTFT synthesis to produce jet fuel [6, ch. 4].

When LTFT is fine-tuned towards jet fuel, the two main products are jet fuel and motor gasoline. Co-LTFT gives a syncrude with the mass product distribution 64%, 32%, and 4% for jet fuel, motor gasoline and other products, respectively. For Fe-LTFT the numbers are 59%, 35%, and 6%. Even though it is tempting to choose the Co-catalyst, A. de Klerk [5, ch. 26] states that it is better to use Fe-catalyst because the Co-catalyst gives motor gasoline that cannot meet the octane rating requirements without additional conditioning. It is also easiest to produce jet fuel from the syncrude from the Fe-LTFT synthesis than any of the other types of Fischer-Tropsch syntheses [6, ch. 4].

2.3 Present day GTL plants

Today, there are five commercial GTL plants operating in the world. They are listed in Table 2.1.

In addition to the existing commercial GTL plants, a lot of R&D is made on small scale GTL plants, where companies such as Greyrock have successfully demonstrated *MicroGTL* plants with capacities of as small as 30 bpd at around 1.5 times the specific investment of large scale plants [14].

Table 2.1: Commercial GTL plants in the world [14].

Plant name	Operator	Geographic location	Plant size [bpd]
Bintulu GTL	Shell	Malaysia	14,700
Escravos GTL	Chevron and Shell	Nigeria	33,000
Oryx GTL	Qatar Petroleum and Sasol	Qatar	34,000
Mossel Bay GTL	PetroSA	South Africa	36,000
Pearl GTL	Qatar Petroleum and Shell	Qatar	140,000

2.4 Danish jet fuel consumption

In the Danish Energy Statistics [15], the aviation sector is divided into three subsectors: domestic, international, and the Danish Defence. In this study, it is the total that is interesting, so the three are combined. In 2017, the total amount of jet fuel consumed by these sectors was 43.6 PJ [15]. The DEA predicts an increased demand for jet fuel of 8% in 2030 compared to that in 2017 [16]. Since it is only 8%, it is considered not to affect the results significantly in regards to the economy with potential *economy of scale* effects to use the numbers from 2017. The efficiency of the plant is also not considered to change by an increase in plant size, so the analysis can be based on the numbers from 2017 without significant error.

2.5 Biogas and biomethane potentials in Denmark

The feed for the GTL plants will in this analysis be biogas and biomethane. To show the Danish potentials of these energy carriers, three numbers are used for both gases. They are the current energy from biogas and biomethane (from 2017) [15], the realistic potential in 2035 [17], and a complete Danish potential based on a meta-study on the subject [18]. The total Danish potential of biomethane does not appear in the meta-study, but assuming a composition of 60% CH₄ and 40% CO₂, methanation can convert the CO₂ to CH₄ and water. The numbers appear in Table 2.2.

Table 2.2: Biogas and biomethane potentials in Denmark.

	Biogas [PJ]	Biomethane [PJ]
2017 levels [15]	11.2	5.1
2035 realistic levels [17]	50	80
Total Danish potentials [18]	100	166.7

2.6 District heating in Denmark

With the supposed generation of recoverable heat from GTL plants, a few key numbers of the Danish district heating system are highlighted here. The yearly district heating production profile of 2014 from Fjernvarme Fyn [19], the utility company in Odense, has been acquired. For this city, the total production in this year was 8.6 PJ. In Copenhagen, the biggest city of Denmark, the total district heating production from the utility company was 16.4 PJ in 2014. Other cities discussed in this study are Aalborg, Esbjerg, Silkeborg, and Vejle with yearly productions of 6.2 PJ, 3.4 PJ, 1.3 PJ, and 0.7 PJ, respectively [20]. In 2017, the total Danish district heating production was around 135 PJ [15], which gives an estimated annual base load in Denmark of 27.2 PJ based on the annual profile of Fjernvarme Fyn.¹

2.7 Economic prerequisites

Although the Fischer-Tropsch technology can be traced back to the first half of the 20th century, there are just five commercial GTL plants around the world today as described in Section 2.3. From the literature on the topic, overall capital expenses (CAPEX) and operational expenses (OPEX) are found and used in the economic analysis. Even though, these costs are related to setups that will not be one-to-one to a setup in Denmark, it is not considered to add more precision to the analysis to manipulate these numbers. This is due to the great uncertainty that is already related to them [5, ch. 1]. An example of this uncertainty is the GTL plant Escravos in Nigeria that was just below ten times as expensive as the cheapest installed plant, Oryx, in Qatar when they are compared on the specific investment [21]. Economy of scale is also said to have an impact on the costs of the plants [6, ch.7]. The in Section 2.3 mentioned R&D in small scale plants, however, shows that the CAPEX may only be around 1.5 times greater compared to large scale commercial plants [14]. The cost breakdowns of CAPEX and OPEX are described below.

2.7.1 Composition of CAPEX and OPEX

Figure 2.1 shows the breakdown of the CAPEX and OPEX. The largest contributor to the CAPEX is clearly the syngas generation that almost comprises half of the total CAPEX. Alternative CAPEX compositions with different shares are also found, where the syngas generation only amounts to around 30% of the total costs [5, 6]. The other large cost of the CAPEX termed *off-sites, utilities, and other units* which refers to product storage tanks, laboratories, buildings, interconnecting, cooling water tower, firefighting water, freshwater, service water, etc. For the OPEX in Figure

¹See the calculation in Appendix F.

2.1b, the largest cost is *catalysts and chemicals* mostly used for the syngas generation which is around 1/3 of the OPEX. Apart from that, the other parts of the OPEX are mostly related to personnel, general administration, and mechanical maintenance of the plant [6]. The feed cost is not included in the OPEX here.

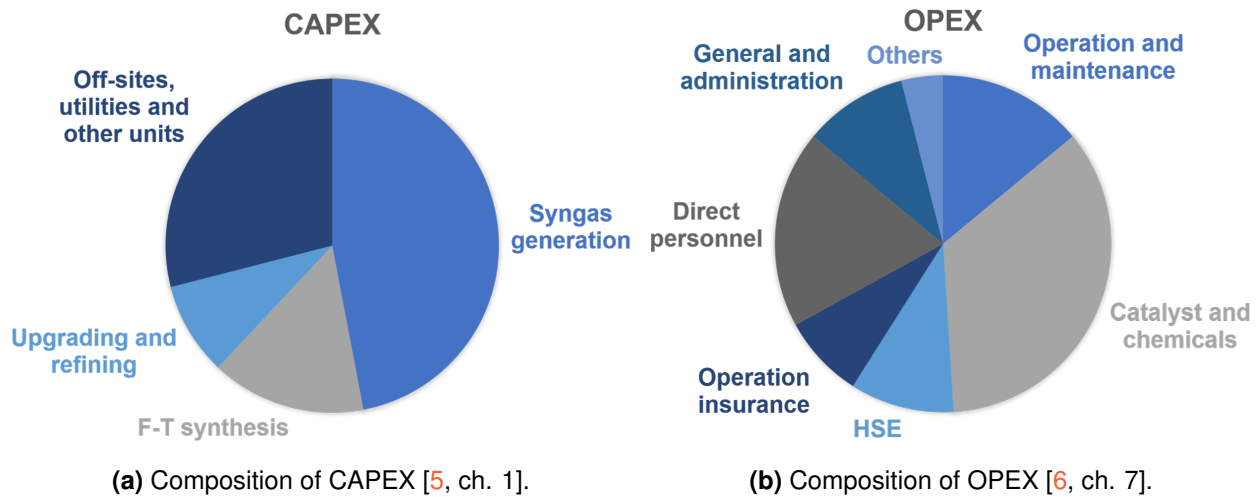


Figure 2.1: Price compositions for CAPEX and OPEX (variable and fixed) used in the analysis. The numbers for the composition are taken directly from the sources. Costs are for a GTL plant with syngas production, Fischer-Tropsch synthesis, and syncrude refining.

2.7.2 Biogas and biomethane prices and subsidies

The typical cost of biogas is between 130-142 DKK/GJ. For traditionally upgraded biogas, the cost is typically between 154-166 DKK/GJ [22]. The cost of methanation is between 162-189 DKK/GJ [18]. For the calculation of the net present value (NPV), the average in the intervals will be used, i.e. 137 DKK/GJ for biogas and 175.5 DKK/GJ for biomethane that comes from upgraded biogas with methanation [22].

Today, different subsidy schemes cover the biogas and biomethane industry. Today's subsidies are divided into three parts. (1) a basic amount which is index regulated, (2) a supplement that depends on the natural gas price, and (3) a supplement that is phased out in 2020. For traditional upgrading, the biomethane can be expected to be subsidised by approximately 105 DKK/GJ, and when biogas is used for processes it will be around 65 DKK/GJ in 2020. Presently, methane from methanation of CO₂ is not subsidised [17].

2.7.3 Price for CO₂ extraction

As will be shown, the performance of the GTL plant increases if CO₂ co-feed is used. CO₂ capture can be either direct air capture (DAC) or point source capture. The carbon capture technology is still relatively young, so the cost per ton of captured CO₂ is conservatively chosen to be 600 DKK/ton, although, it seems possible to achieve costs around half of this with future improvements of the technology [23, 24].

2.7.4 Jet fuel and gasoline prices

For the business economic analysis, the willingness-to-pay (WTP) for bio jet fuel is assumed to be 250 DKK/GJ based on a statement from *Henrik Wenzel* [25] in a recent newsletter. For the bio-gasoline, the DEA projections for conventional gasoline as given in the socio-economic calculation prerequisites is used [26], as consumers of this product may not be willing to pay extra for biogasoline. This is done mainly due to the energy tax on *gasoline equivalents* of 128.1 DKK/GJ stated in the *Mineralolieafgiftsloven* [27] that is imposed even on green fuels, which will make the WTP highly reliant on the image value of the product, whereas bio jet fuel gives the airlines an additional saving on avoided CO₂ quotas [28].

For reference, the conventional jet fuel average price for 2019 is \$80.8/bbl equal to 100 DKK/GJ² taken from the International Air Transport Association, IATA [29]. The conventional gasoline price is almost the same [26]. According to the reference (best assessment) in the Annual Energy Outlook 2019 by the U.S. Energy Information Administration, EIA, the oil prices are not expected to change a lot, although, their high estimate almost doubles the reference [30].

2.7.5 CO₂ quota price

The present-day CO₂ quota price is already a lot higher than what is predicted in the calculation prerequisites of the DEA, and higher than their expectations for the next 20 years [26]. Today's price of 200 DKK/ton [31]³ will, therefore, be used in the analysis and not the DEA projections. The CO₂ emission coefficients for jet fuel and gasoline are 73 kg/GJ [15]. The CO₂ quota price is used for the socioeconomic analysis as the cost of the social harm.

²As of 19/5-2019 with an exchange rate of 6.6936 DKK/USD

³As of 12/5-2019

2.7.6 Corporation tax

The corporation tax in Denmark is 22%. It is to be paid on the profit of a corporation when all expenses, including depreciation of assets, are subtracted from the revenues [32]. This is included in the business economic NPV calculation.

2.7.7 District heating remuneration

As there will be a certain amount of recoverable heat to be used for district heating, the revenue from this should be included in the analysis. In 2015, the weighted average remuneration for the sales of district heating in Denmark was 141 DKK/MWh equal to 39.17 DKK/GJ in 2015 prices [33].

3 Methods

This section describes the simplifications that are made in relation to the models and the economic methods used in the analysis.

3.1 Simplifications

In this study, simplifications are made regarding the complicated reactions and product compositions related to the production of jet fuel in GTL plants, so they are reduced to as simple components as possible without changing the results too much. For example, the biomethane is said to be perfectly clean methane, the biogas is a perfectly clean gaseous mixture of methane and CO_2 in the ratio 60:40 [34] and also the water is pure H_2O .

For the products, the jet fuel is considered to be composed of undecane with the chemical formula $\text{C}_{11}\text{H}_{24}$ and the gasoline that is the main by-product from the analysed refining method is octane with the chemical formula C_8H_{18} . The fuel gases propane and butane, C_3H_8 and C_4H_{10} , describe the other by-products from the process. This is reasonable as the finding in Section 4.1 shows that the heat duty does not differ significantly from alkane to alkane when it is divided by the carbon atoms in the respective alkane.

Jet fuel is taken to be undecane since the average formula for jet fuel is $\text{C}_{11}\text{H}_{21}$ which is similar to that of undecane on the number of C-atoms [35]. For motor, gasoline octane is used since iso-octane is used as the reference for gasoline when the resistance to self-ignition is benchmarked [36].

Jet fuel has an LHV around 43.8 MJ/kg [15] where undecane has a slightly higher LHV of 44.2 [37]. Furthermore, undecane has a higher freezing point of -25.6°C [38] against the required maximum freezing temperature of -47°C for jet fuel of the type JET A as stated in the UK specification DEF STAN 91-91 [39]. It is, therefore, not an option to use undecane as jet fuel in reality. The effect of using a component with a higher LHV in the calculations is that the heat generation is smaller when the component is a product in the chemical reaction. If it is a reactant it will be larger. The same goes for the gasoline that also has an LHV of 43.8 MJ/kg [15] against that of octane of 44.4 MJ/kg [37]. This is not regarded as a big source of error.

With these product simplifications, the optimal $\text{H}_2:\text{CO}$ ratio is 2.1062 based on the product yield for Fe-LTFT described in Section 2.2.2.⁴ This is similar to what is described in the literature, where it is typically around 2 [5, ch. 1] and e.g. 2.15 at the Bintulu GTL plant in Malaysia [5, ch. 11].

⁴For calculation see Appendix B.

On this basis, the Aspen Plus model is made with the reformer and the reactor as R-GIBBS reactors where the possible products are limited so as to include the effect of the catalysts. A discrepancy between the model and real world plants will mainly be due to this. In Section 4, these methods are verified for the analysis and a description of the setup of the plant is given.

3.2 Economy

The CAPEX and OPEX that are used are found in USD in 2010 and 2011 prices, respectively. They are converted to DKK according to the average exchange rate in 2010 and 2011 [40, 41] and then converted to present value using the net price index from January each year [42]. The index regulated prices are calculated as

$$\text{value}_{2019} = \text{value}_{\text{past}} \cdot \frac{\text{index}_{2019}}{\text{index}_{\text{past}}} \quad (1)$$

3.2.1 Levelised cost of energy

The levelised cost of energy (LCOE) is calculated from the economic inputs and the efficiency values from models made in Aspen Plus where all the costs are converted into a price per GJ.⁵ This requires amortisation of the CAPEX. An interest rate of 12% and a depreciation period (equal to the total lifetime of the plant) of 25 years are used [6, ch. 7]. The interest rate is set this high because the technology is still not entirely commercialised [18], and is therefore considered to pose a greater risk for investors. The formula used for calculating the annualised CAPEX is

$$A = P \cdot \frac{i(1+i)^N}{(1+i)^N - 1} \quad (2)$$

where A is the annuity, P the present cost, i the interest rate, and N the depreciation period in years.

3.2.2 Net present value

The NPV will be calculated with the same interest rate as the LCOE and the same lifetime, i.e. 12% and 25 years. It is calculated by the formula

$$\text{NPV} = -C_0 + \sum_{t=1}^N \frac{C_t}{(1+i)^t} \quad (3)$$

⁵See in Appendix G for the calculations.

where C_0 is the investment in the zeroth year, t the year, N the lifetime, C_t the net annual cash flow, and i the interest rate.

3.2.3 Socioeconomics

The guidelines from the Danish Ministry of Finance [43] are followed for the socioeconomic calculation. They propose an interest rate of 4% and a net tax factor of 1.325 imposed on all costs where all taxes and subsidies are removed. The CO₂ quota price is the basis for the social harm that the CO₂ emissions impose on society. Equation 3 is the same for the social NPV, although, the monetizable harm and benefit should be included in the net annual cash flow.

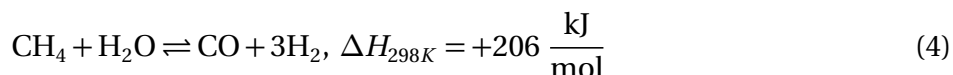
4 Aspen Plus

The already mentioned commercial GTL plants are located all around the globe. However, when investigating the implementation of such a plant in the Danish energy system, it is of the utmost importance to know how much recoverable heat is generated to know if it can be used in the district heating system. At the same time, different types of feed are investigated which will undoubtedly have an impact on the performance. A large part of this study has, therefore, been to build a model in Aspen Plus where the performance in relation to the district heating and product efficiency of GTL plants for jet fuel production can be evaluated. To verify the model, simulation results are compared with stoichiometric calculations made in Excel.

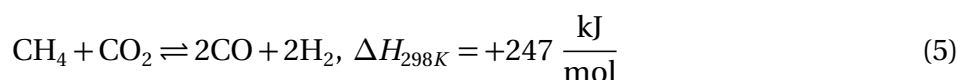
4.1 Stoichiometric calculations vs. simulations

Stoichiometric calculations are made under standard conditions which are a temperature of 298 K (25 °C) and a pressure of 1 atm., although with steam at 398 K (125 °C) but at the same pressure. Using steam in stoichiometric calculations is equivalent to comparing the LHV on both sides of the reaction arrow. This is one of two methods for calculating the enthalpy change of the reactions and this is done in the following. The other method is by evaluation the difference in the enthalpy of formations of the reactants and the products [44, ch. 15].

When methane is converted into syngas using steam without any co-feed of CO₂, the stoichiometric reaction is



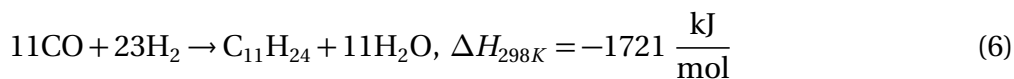
and when CO₂ is added the additional reaction is



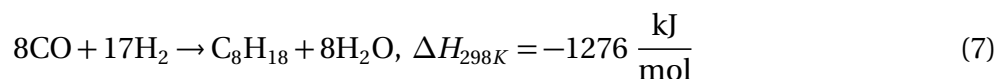
The positive changes in enthalpy in the above reactions indicate that both of the reactions are endothermic and require heat. Endothermic reactions also typically require high temperatures, why standard conditions will not be present in reality [45]. Furthermore, some side reactions will happen, and some fraction of the reactants will pass through the reactor unreacted [44].

After the reforming, let alone the syngas cleaning and conditioning, the syngas goes through the Fischer-Tropsch synthesis where the H₂ and CO are converted into hydrocarbons. The products of the synthesis are alkenes, alkanes, alcohols, carbonyls, carboxylic acids, and water gas shift products. The current model evaluates the conversion into alkanes only, where the reaction into

jet fuel and gasoline are



and



Both of the reactions are exothermic and heat must be removed from the reactor [5, ch. 3, 4].

Notice that the energy per C-atom is nearly unchanged (156.5 kJ/mol/C-atom in Equation 6 and 159.5 kJ/mol/C-atom in Equation 7), so in the calculations used to compare the results from Aspen Plus, only Equation 6 is used. The optimal ratio of the reactants in this reaction is 2.09.⁶ This ratio is neither obtained from Equation 4 nor 5. When $\text{H}_2:\text{CO} > 2.09$ there is a surplus of hydrogen which lowers the product efficiency, and when $\text{H}_2:\text{CO} < 2.09$ CO_2 is formed which lowers the carbon efficiency. In four consecutive calculations, the required heat input to the reformer and heat output from Fischer-Tropsch synthesis are calculated with varying $\text{CH}_4:\text{CO}_2$ ratios. The ratios evaluated are 1:0, 3:1, 3:2, and 1:1. Note that 3:1 gives a $\text{H}_2:\text{CO}$ ratio of 2, so this is slightly modified to 77:23 (but still called 3:1) which yields the optimal ratio of 2.09 $\text{H}_2:\text{CO}$. All the calculations are made to give exactly one mole of $\text{C}_{11}\text{H}_{24}$. The results are compared with results from Aspen Plus where the same operating conditions are used to simulate the two simplified steps from gas to liquids in R-GIBBS reactors where the products are limited to the discussed ones.⁷

Figure 4.1 shows the results of the comparison. The greatest relative difference between the calculations and the Aspen Plus simulations is 0.7%. The difference is greatest when the $\text{CH}_4:\text{CO}_2$ ratio is 1. The optimal $\text{H}_2:\text{CO}$ ratio of 3:1 is also the one requiring the least amount of energy to produce one mole of jet fuel.

4.2 Aspen Plus model setup

The model is optimised to remove the need for fuel-fired pre-heating of the inputs to the SMR or SDR unit.⁸ The model setup in Figure 4.2 can be followed in the following description. The reference temperature is set to 25 °C. In order to pre-heat the water and the feed to the reformer, several heat exchangers are installed. For the water, there are an economiser (1), an evaporator (2), and a superheater (3). The hot stream in these three heat exchangers is the flue gas from the process heater for the reformer. The flue gas first goes to the superheater (3), then the evaporator (2), and at last the economiser (1) to make the most use of the high temperature. After the

⁶For the calculation see Appendix B

⁷Find the Excel file in Appendix L.

⁸See Appendix K for a detailed description

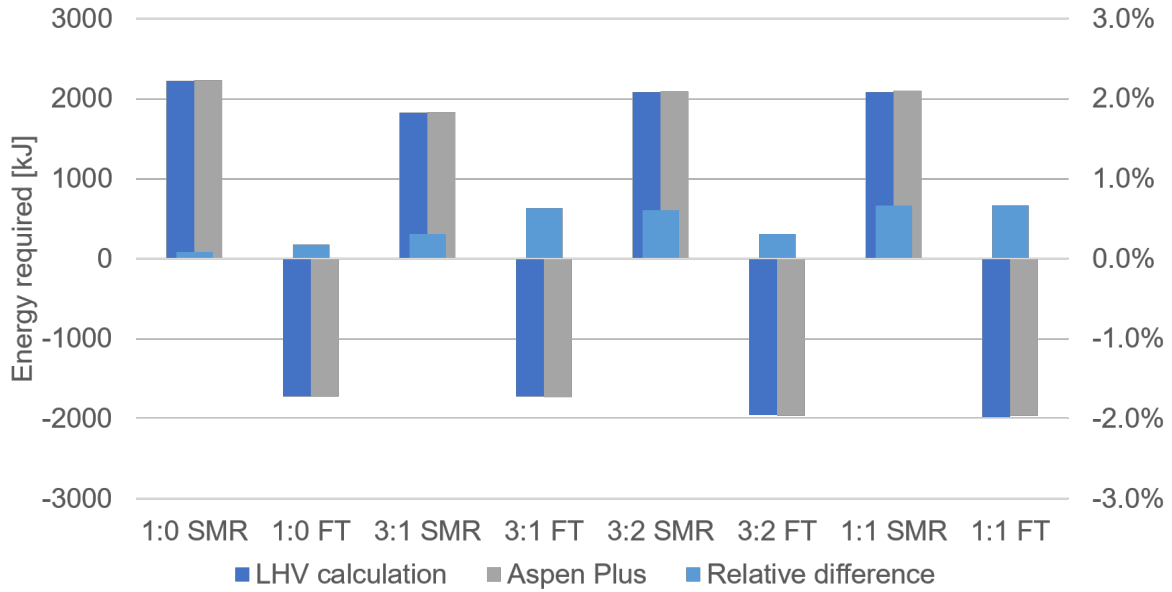


Figure 4.1: Comparison between calculations from LHV and simulations results from Aspen Plus. *SMR* is energy required by the steam reformer and *FT* is the energy requirements for the Fischer-Tropsch synthesis. *LHV calculation* and *Aspen Plus* relate to the left y-axis. The *Relative difference* relates to the right y-axis.

flue gas has pre-heated the water, it is led to one of two superheaters (4) for the gaseous feed (CH_4 and CO_2). The second of the superheaters for the feed (5) is heated by the syngas. The syngas is also used for heating the syncrude (6) to 350°C which is the temperature at which the separation, the first step in refining process, takes place [46]. Furthermore, the hot syngas is also circulated, so it can pre-heat itself (7) before the Fischer-Tropsch synthesis as the syngas conditioning previous to the synthesis requires low temperatures [5, ch. 1]. The model only has the reforming and the synthesis and stops right before the refinery, however, the inefficiencies in the refining process are included (8) in the model. Consequently, the end products are is if it was a complete GTL plant.

The $\text{H}_2\text{O}:\text{CH}_4$ molar ratio in the steam reformer is set to 2.5. The operating pressure is 20 bar and the temperature is 900°C [5, 7], however, a pre-study is made to investigate the effect of the pressure in the reformer as findings from other studies have shown that the syngas production performs better at lower pressures [47, 45]. The Fischer-Tropsch synthesis in the model is simplified and split into three parts, where the syngas going into each part is split according to the mass fraction of the products.⁹ Operating temperature and pressure of the Fischer-Tropsch synthesis are 200°C and 10 bar.

District heating operating conditions are approximated according to Danish systems with a temperature to the GTL plant at 40°C and from the plant at 70°C and at a pressure of 5 bar [48].

⁹59% jet fuel, 35% gasoline, and 6% other products [5, ch. 26].

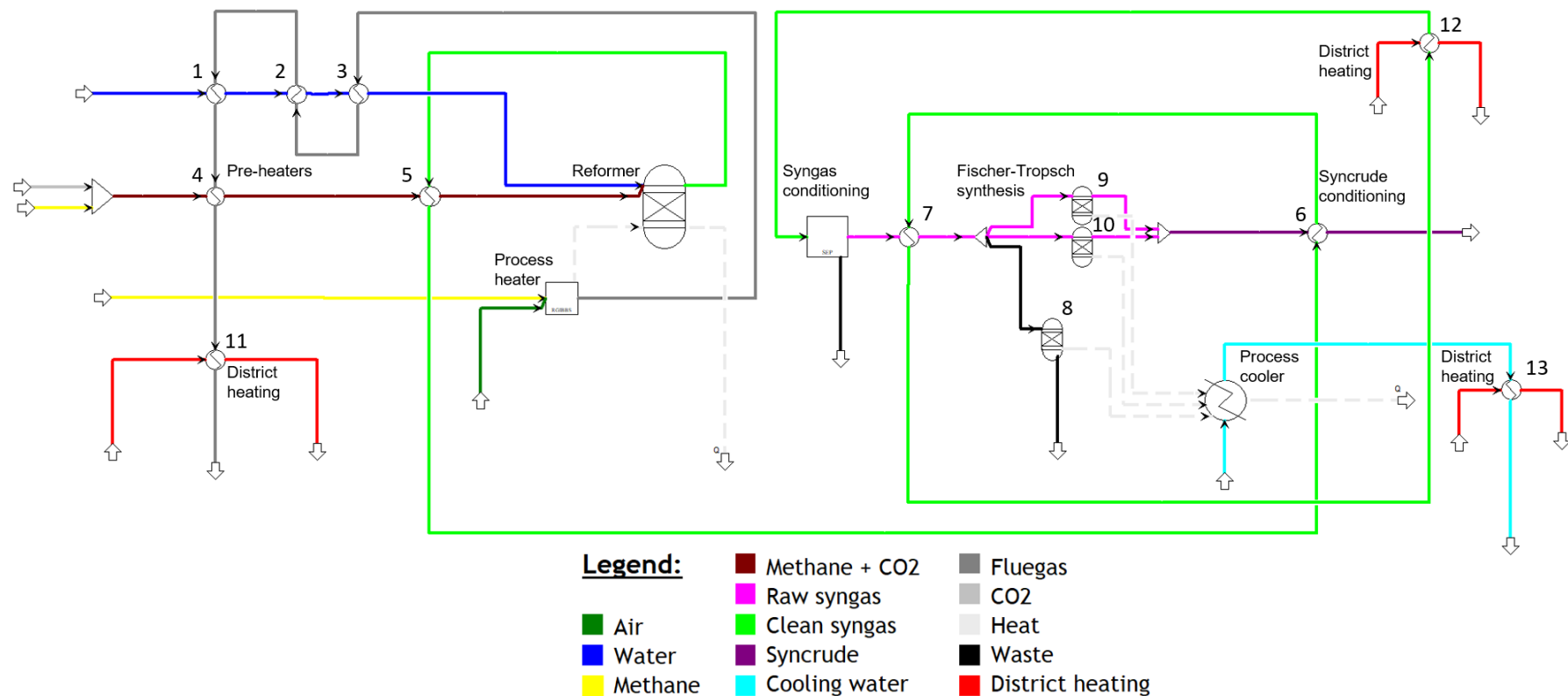


Figure 4.2: Aspen Plus model setup used in calculations. The configuration of the main blocks in Aspen Plus are listed in Table A2 in Appendix A. 1, 2, and 3 are pre-heaters for the steam to the reformer, 4 and 5 are pre-heaters for the feed, 6 is a pre-heater for the syncrude and 7 is one for the syngas. 8 is the part of the Fischer-Tropsch reactor that forms the hydrocarbons that are referred to as *inefficiencies* e.g., fuel gases, unrecoverable organics, and petrochemicals, not included in the analysis on an energy basis. 9 and 10 represents the synthesis into jet fuel and gasoline. 11, 12, and 13 are the heat exchangers that are used to calculate the amount of recoverable heat. The main blocks in the figure have names associated with them.

5 Pre-study

Before defining the scenarios, it is investigated how the GTL model performs when the CH_4 to CO_2 ratio in the feed is varied. In Aspen Plus, the described model is simulated where the ratio is varied from 100:0 to 45:55 and the efficiencies are compared.

As mentioned in Section 4.2, the impact of changing the pressure from 20 bar to 5 bar in the reformer is tested. All other parameters are kept constant. The results are as predicted; the product efficiency is higher at lower pressures. The difference, though, is small and the 5 bar reformer performs only around one percentage point better than the 20 bar reformer when the difference is greatest. With this in mind and knowing that the low-pressure reformers are unconventional, they are disregarded in this study [5, 6, 7, 49]. A graphical comparison can be seen in Figure 5.1.

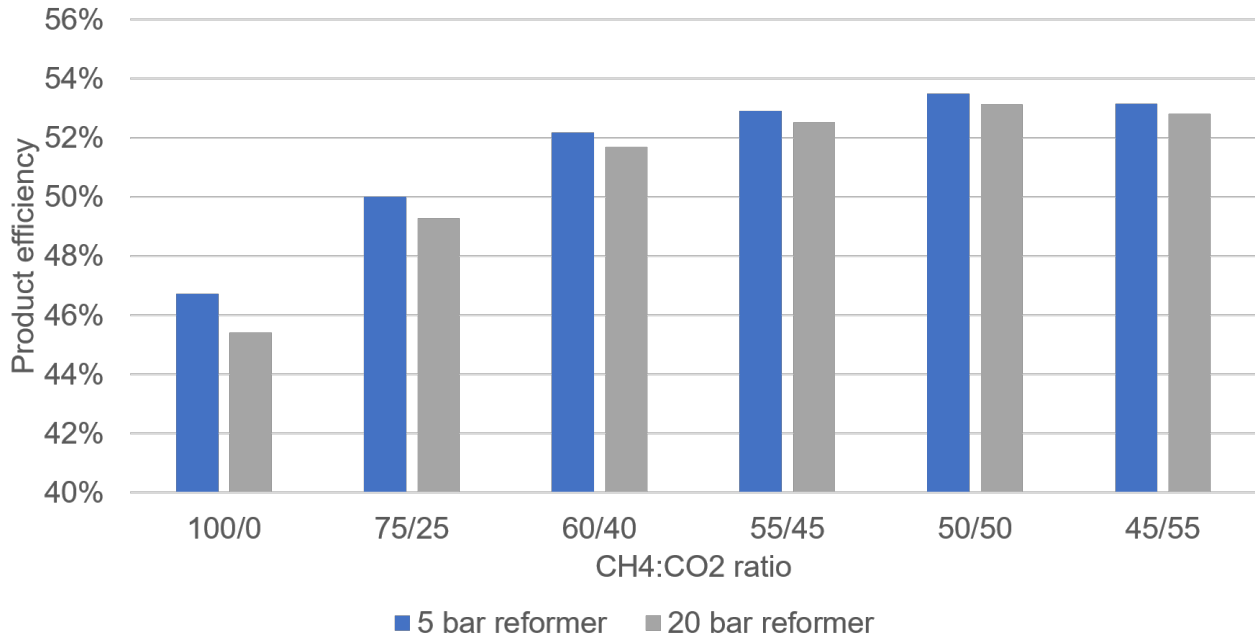


Figure 5.1: Product efficiency over a range of CH_4 : CO_2 ratios in a 5 bar and a 20 bar reformer. The difference is not significant.

To get the full picture of the different setups, the district heating efficiency should also be included. The gas loops of the tail gases were found to have a huge impact on the setups, as the ones that perform the worst has the greatest surplus of hydrogen which is wasted if the gas loops are not used. For the setups with the 20 bar reformer and gas loop the results are in Figure 5.2. In this analysis, the tail gas is the excess hydrogen and unreacted methane from the reformer and the hydrogen that is formed in the simplified Fischer-Tropsch reactor. Notice that the product efficiency and the total efficiency both increase as the amount of CO_2 is increased until a certain point while the opposite is true for the district heating efficiency. The district heating efficiency

is only the potential if all recoverable heat is used. The results from the simulation are in line with findings from other studies [18, 50, 7].

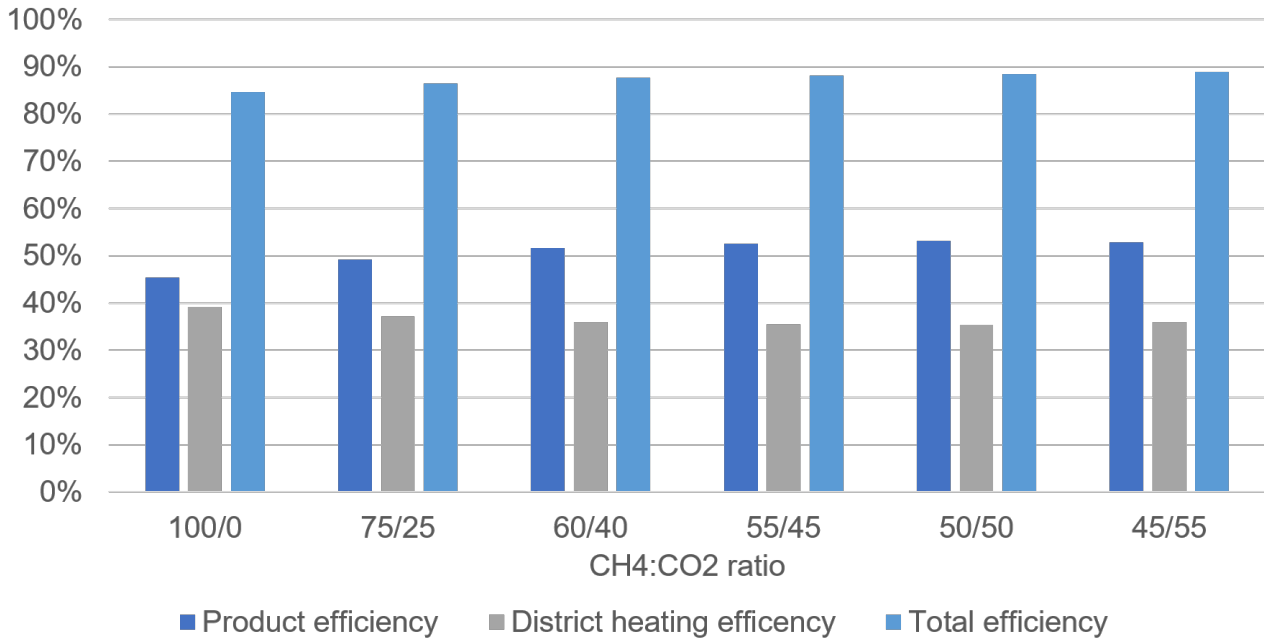


Figure 5.2: Product, district heating, and total efficiencies depending on the amount of co-feed CO₂.

By using *user defined* blocks in Aspen Plus, the CH₄:CO₂ ratio can be adjusted to give the desired H₂:CO ratio of 2.1062 which is optimal for this model as described in Section 3.1. To achieve this H₂:CO ratio in the syngas, the feed CH₄:CO₂ ratio should be 51:49 which is only slightly greater than 1. It is not equal to the results from the stoichiometric calculations in Section 4.1 due to inefficiencies caused by the nature of chemical processes related to Gibbs energy minimisation resulting in e.g. methane slip and CO₂ formation in the reformer. The carbon efficiency of the reformer is, thereby, 64% with an optimal CH₄:CO₂ ratio. The results of the entire optimised GTL conversion are a product efficiency of 53%, a district heating efficiency of 35%, and carbon efficiency of 57%. In Appendix C, a graph focusing on the product efficiency around the optimum can be found.

Furthermore, the impact of the pumps, compressors, and separation in the steps affect the total efficiency of the system. This is tested, and the results can be found in Appendix C.1. They do not constitute a significant energy consumption in the overall GTL process. This justifies the simplification of neglecting the energy consumption of the pumps and compressors in the system.

6 Scenarios

Based on the findings regarding the performance of the GTL plant found in Section 5, three scenarios for Fischer-Tropsch jet fuel production in Denmark is proposed. Each of the scenarios have some expected advantages and disadvantages which are described in the respective sections. The feed to the GTL plants in the three scenarios are biogas, biomethane, and biomethane with additional CO₂ co-feed. In all scenarios, the plants are assumed to run year-round. The scenarios are described below.

6.1 GTL, Biogas

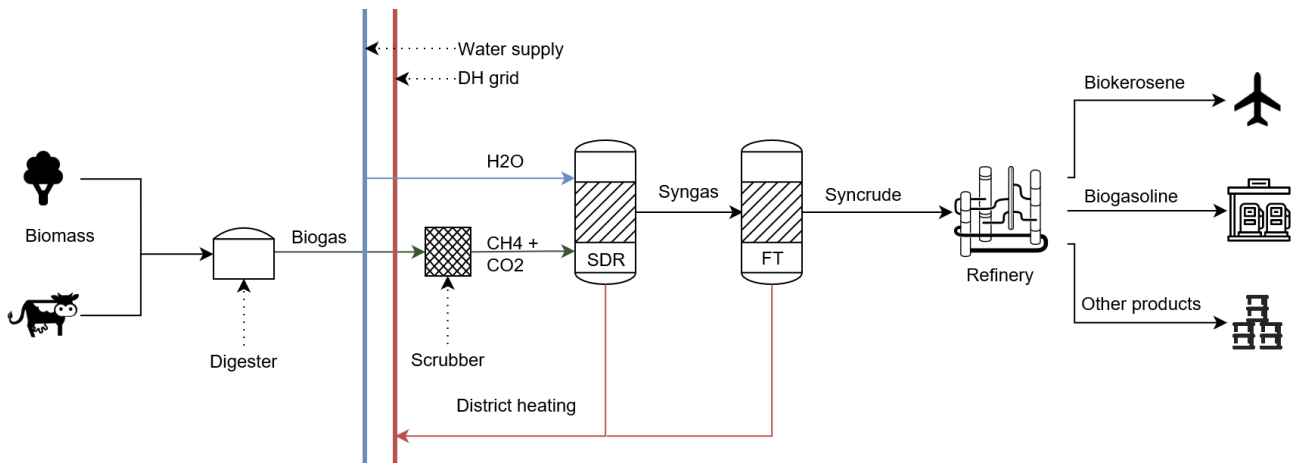


Figure 6.1: System diagram of the scenario where the feed is biogas with a CH₄:CO₂ ratio of 3:2.

This scenario uses biogas that is cleaned for impurities such as hydrogen sulfide (H₂S). This leaves the biogas with a composition of 60% CH₄ and 40% CO₂ [34]. Since biogas is used directly in this scenario and since there is not a grid for biogas in Denmark, the GTL conversion should be decentralised close to where the biogas is produced. In reality, this can be done in two ways; either the entire GTL conversion takes place decentralised or the syncrude that is produced decentrally can be refined at large scale refineries. A system visualisation of the scenario is shown in Figure 6.1. Advantages and disadvantages are listed below.

Advantages:

- Already a high CH₄:CO₂ ratio.
- High product efficiency.
- CO₂ is already present and is not extracted from a secondary source.

- Smaller plants give better utilisation of the surplus heat if recoverable heat is huge.
- Lowest feed price.

Disadvantages:

- No economy of scale.
- Small scale GTL plants are more immature.
- Lowered energy potential due to the lack of methanation of biogas.
- Decentral plants may be too far from district heating areas.
- Many small plants require many investors.
- Biogas upgrading is already widely used [15].

6.2 GTL, Biomethane

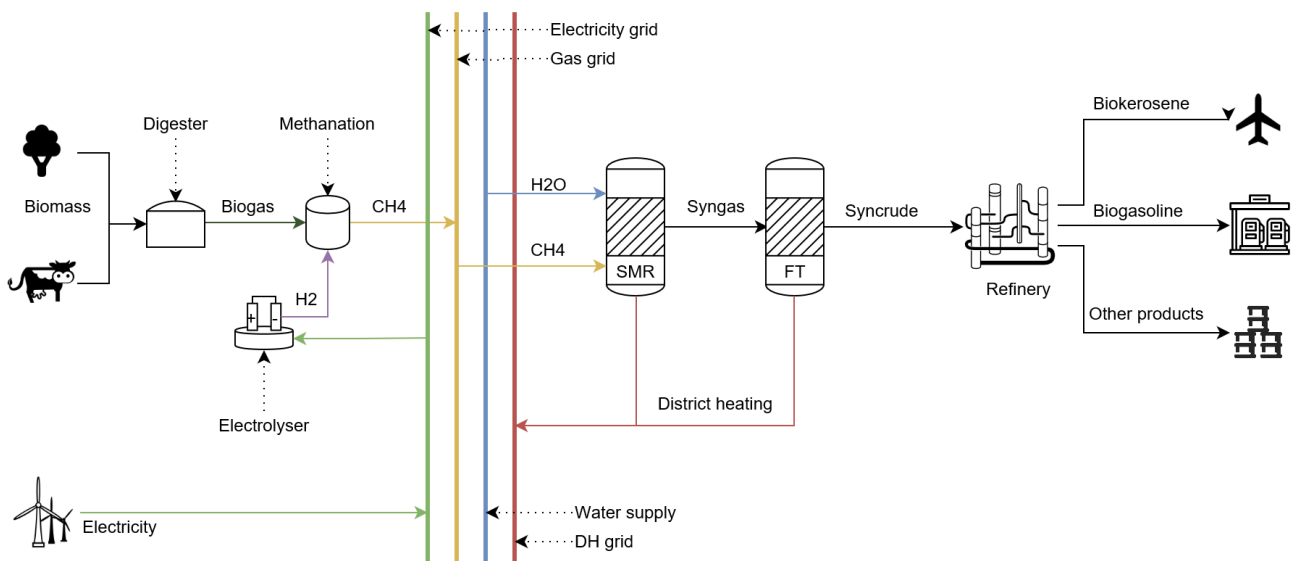


Figure 6.2: System diagram of the scenario where the feed is biomethane which is 100% CH₄.

This scenario uses biomethane which is assumed to be 100% CH₄. The biomethane can be transported in the natural gas grid that is already installed in Denmark and the GTL plant can be centralised. As this scenario will have the lowest product efficiency, it will also be the one that requires the greatest methane input to satisfy the jet fuel demand. A system visualisation of the scenario is shown in Figure 6.2. Advantages and disadvantages are listed below.

Advantages:

- Can use the current natural gas grid.

- Large central plants can take advantage of economy of scale.
- CO₂ is not added and is not extracted from a secondary source.
- High energy potential due to methanation of biogas.
- Central plants can be placed in large district heating areas.
- Biogas upgrading is already widely used [15].

Disadvantages:

- Lowest product efficiency.
- Highest methane input required to satisfy the jet fuel demand.
- Recoverable heat from central plants may be too large for the district heating system.

6.3 GTL, Biomethane with CO₂ co-feed

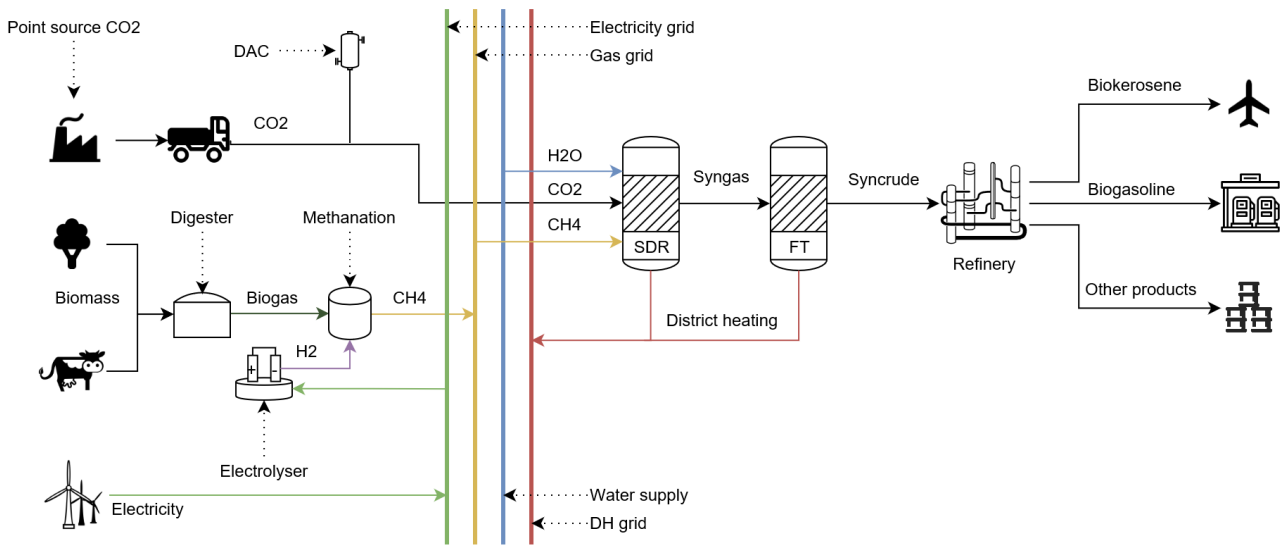


Figure 6.3: System diagram of the scenario where the feed is biomethane with CO₂ which has the optimal ratio of CH₄:CO₂ of approximately 51:49.

This scenario stands out from the other two scenarios by the addition of CO₂. This makes it possible to optimise the GTL conversion by adjusting the CH₄:CO₂ ratio to 51:49. The CO₂ can come from two types of secondary sources: either by point source capture or DAC. Point source capture can either originate from factories or it can be recaptured from the GTL plant itself. This scenario can also make use of the natural gas grid, so the GTL plant can be centralised. A system visualisation of the scenario is shown in Figure 6.3. Advantages and disadvantages are listed below.

Advantages:

- Possibility to adjust the $\text{CH}_4:\text{CO}_2$ ratio to its optimum.
- Can use the current natural gas grid.
- Large central plants can take advantage of economy of scale.
- High energy potential due to methanation of biogas.
- Lowest methane input required due to optimisation.
- Central plants can be placed in large district heating areas.
- Biogas upgrading is already widely used [15].

Disadvantages:

- Carbon capture technology should be used or CO_2 should be bought from other factories which may be more expensive.
- Recoverable heat from central plants may be too large for the district heating system.

7 Analysis and discussion

With the models scaled after the Danish jet fuel demand, the three scenarios give the outputs shown in the Sankey diagrams in Figures 7.1, 7.2, and 7.3. The actual utilisation of the recoverable heat depends a lot on the setup and location of the GTL plant(s) and will be analysed in Section 7.2.2 and discussed further in Section 7.2.3. Appendix M shows the extracted results from Aspen Plus to Excel.

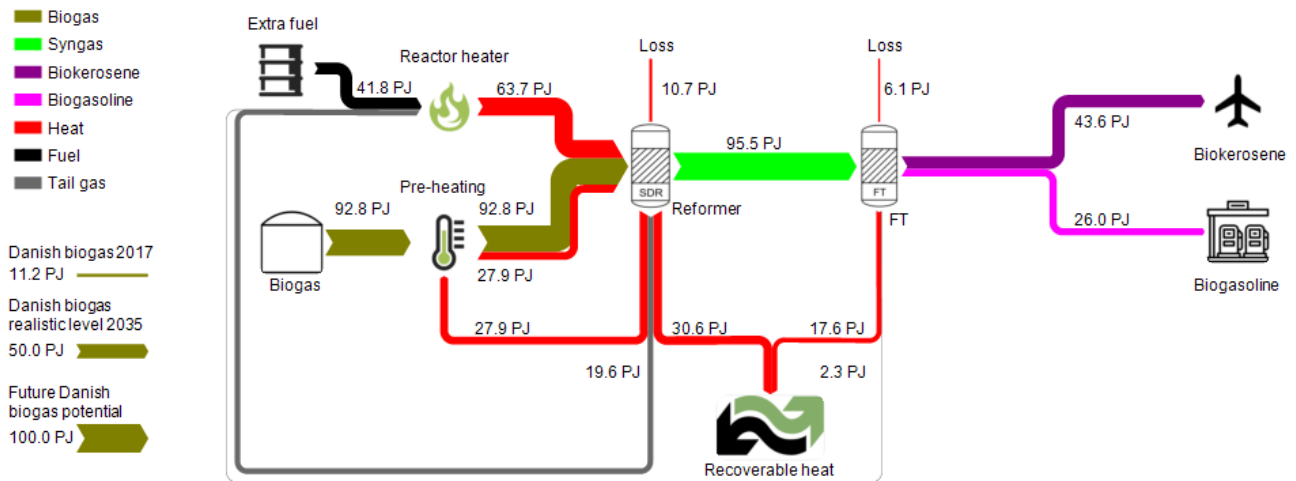


Figure 7.1: Sankey diagram of the biogas scenario.

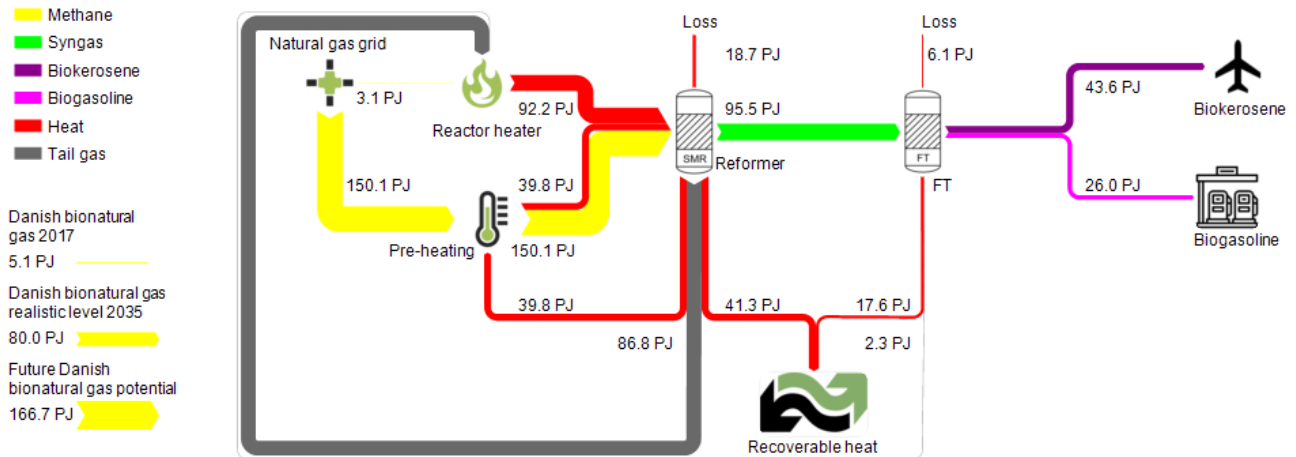


Figure 7.2: Sankey diagram of the biomethane scenario.

The efficiencies shown in Table 7.1 are the same as in the pre-study since the only difference is the scaling of the models. Product and district heating efficiencies are the ratios between the energy in the liquid products and the district heating output to the total external energy to the

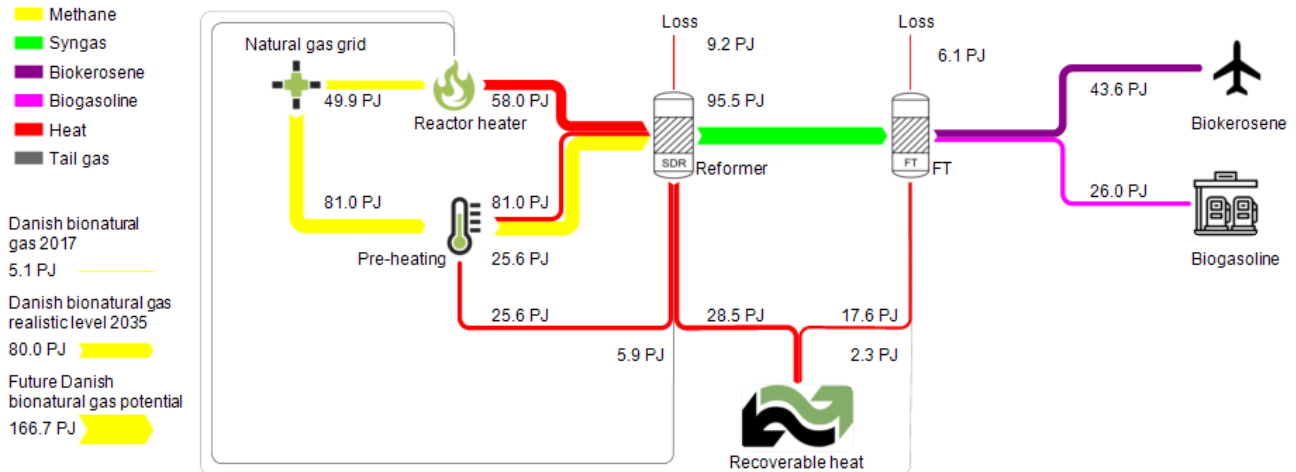


Figure 7.3: Sankey diagram of the biomethane and CO₂ scenario.

process in the Sankeys. These are used in the economic analysis when the performance of the plant has an impact on the costs, i.e. the different feed costs. Notice that the carbon efficiency decreases with increasing product efficiency. This is due to the fact that carbon in the co-fed CO₂ is not converted to products at the same rate as the methane. The results from the simulations resembles the numbers found from other studies in spite of the simplifications [18, 50, 7].

Table 7.1: Product, district heating, total, and carbon efficiencies for the three scenarios.

Scenario	η , product	η , district heating	η , total	η , carbon
Biogas	52%	36%	88%	57%
Biomethane	45%	38%	84%	59%
Biomethane with CO ₂	53%	35%	88%	56%

The scenario with biomethane and co-feed CO₂ has the highest product efficiency as well as total efficiency, and is henceforth also referred to as the *optimal scenario*. The second best is the one using biogas, and the scenario with the poorest performance is the one with biomethane as feed.

7.1 Biogas and biomethane availability

All scenarios require significantly more biogas or biomethane compared to current productions as can be seen from Table 7.2 and the Sankeys. Theoretically, the needs are smaller than the total estimated Danish potential of biogas and biomethane, though. For the biogas scenario, however, it should be noticed that it keeps just below the total Danish biogas potential in regards to the biogas that is required for the feed to be converted, but the process still requires additional

energy for keeping the reformer running (*Extra fuel* in Figure 7.1) where the biogas potential is no longer sufficient.

Table 7.2: Biogas and biomethane availability with the demand for each of the three scenarios.

	Biogas	Biomethane	Biomethane and CO₂
Annual demand	92.8 (+41.8) PJ	153.2 PJ	130.9 PJ
Realistic level in 2035	50 PJ	80 PJ	80 PJ
Total Danish potential	100 PJ	166.7 PJ	166.7 PJ

Under the premise that the aviation industry must change its fuel to Fischer-Tropsch synthesised biokerosene, five of the possible solutions — which can be combined — for the biogas or biomethane shortage are:

1. Reducing fuel consumption — either by flying less or improving fuel efficiency.
2. Investing in harvesting the total Danish biomethane potential.
3. Investing in R&D of the SOEC co-electrolysis technology to lower the need for methane.
4. Importing biokerosene from other countries.
5. Methanation of captured CO₂.

The first point is a possibility since taxes on jet fuel and, thereby, flight tickets are expected to lower the number of flight departures, as for example stated in a newly leaked EU report [51]. However, even if it is lowered by 10% as stated in the news article, the demand will not change significantly to solve the problems. For the second option it can easily be imagined that the marginal cost of extracting energy from biomass will increase as new biomass conversion technologies are to be used, e.g. the conversion of straw which is possible, but still with some technical issues [22]. The third option, as already mentioned in Section 2.2.1, of including co-electrolysis in the system for syngas generation may be a feasible solution if R&D can lower the cost of this technology. The fourth option may also be an alternative, but it only makes sense in the bigger picture if other countries are able to produce more than their own demand. Otherwise, the CO₂ emissions will simply be moved across borders not solving any global emission problems. Finally, methanation of CO₂ is a possibility which combines two of the already discussed technologies. Here, CO₂ must be captured, whereupon methanation with hydrogen forms methane with the CO₂. This is another expensive solution with current technologies.

7.1.1 Fischer-Tropsch jet fuel production in the energy system

If the pathway of Fischer-Tropsch jet fuel production is pushed forward by the Danish government, the industry, and the research institutions, the realistic biogas level in 2035 shown in Table 7.2 might be even higher and closer to the estimated total potential. At the same time, it will require that close to all the entire biogas production is aimed towards Fischer-Tropsch synthesis to make enough bio jet fuel to satisfy the Danish demand by domestic production

Other sectors that today use natural gas should in a zero-emission society also be converted to a fossil-free alternative [52]. In 2017, 32 PJ of natural gas was used for electricity and district heating production, 31 PJ went into the industrial sector, 8 PJ to the commerce and service sector, and 24 PJ was used in households for cooking and individual heating [15]. The sum of these amounts alone are enough to use most of the available biomethane if the consumption figures of today are used. In the future projections, these sectors will reduce their overall natural gas consumption over the years [52], however, it is still not enough to avoid issues on biogas availability.

The road transport sector, i.e. buses and light and heavy duty freight, and the sea transport sector are also expected to be converted to be fuelled by green gases or other biofuels, such as biodiesel or SNG and only to a small extend electricity [3]. With Fischer-Tropsch jet fuel production this should be revisited and instead of gas, it should be investigated whether they can be converted to drive on the biogasoline that is the major by-product from the Fischer-Tropsch jet fuel production. The DEA's energy scenarios [3] also show that most other sectors have several alternative energy sources, whereas the aviation sector is just pictured to have a single alternative to the conventional fuel which is biokerosene. This makes it obvious that this pathway with green Fischer-Tropsch jet fuel should be investigated further if a zero-emission society should be obtained. As mentioned, another solution is to import the biokerosene used for aerial transport. So, if this is less expensive than importing green fuels for the other sectors, 100% jet fuel self-sufficiency may not be the correct solution.

In addition to the sectors that use natural gas today, other industries that today use oil products, should be converted to renewable alternatives, as well. These include, among others, the plastic and steel industries. In Denmark in 2017, after the transport sector, the production sector consumed 36 PJ of oil, which should also be accounted for. Some of this may come from the Fischer-Tropsch process though, as 1.4% of the mass of the refinery products are petrochemicals, that can be used in the production of bioplastics [5, ch. 26]. So, when the entire system is looked upon, more complications arise in relation to the biomass consumption, and this illustrates even more the fact that the overall system should be planned thoroughly when it should be free from fossil fuels.

7.2 Technical analysis

In the following analysis, the technical aspects of the Fischer-Tropsch jet fuel production will be evaluated where the recoverable heat is put in context to the Danish district heating system. Possible locations are discussed as well as the size of the plants.

7.2.1 Size of plant

With a total yearly production of 69.6 PJ of liquid products from the GTL plant, the size of the plant must be around 34,000 bpd.¹⁰ Comparing this to the commercial plants, see Table 2.1, three of the plants have capacities between 33,000 bpd and 36,000 bpd. A one plant solution will, therefore, have a size that has already proved its functionality, and thereby, this is a realistic size. At the same time, R&D in small-scale GTL plants does not exclude the biogas scenario with a lot of distributed production sites. The biogas scenario could for example consist of around 75 small plants around the country where there are already biogas plants, which will give an average plant size of 453 bpd [53, 54]. With the previously mentioned MicroGTL plants of down to 30 bpd, a wide variation in the size of the installed GTL plants is possible depending on the exact plant locations and the biogas plant sizes.

7.2.2 District heating potential

When the total annual consumption of jet fuel is to be produced, the three scenarios generate between 46.1 PJ and 58.9 PJ of recoverable heat per year. In Figure 7.4, an illustrative example of the total recoverable heat compared with the annual heat production from Fjernvarme Fyn in Odense can be seen. The grey area in the figure is not the background but represents the recoverable heat from the optimal scenario. From this, two main conclusions can be drawn. First, the recoverable heat from the GTL plant is constant, whereas the district heating demand fluctuates throughout the year.¹¹ Second, the generated amount of heat is significantly larger than the actual demand in Odense. So, if a single plant with this size is built, not all of the recoverable heat can be used for district heating and must be cooled otherwise, e.g. by a cooling tower.

¹⁰See Appendix D for the calculation

¹¹See the equation for the power in Appendix E.2.

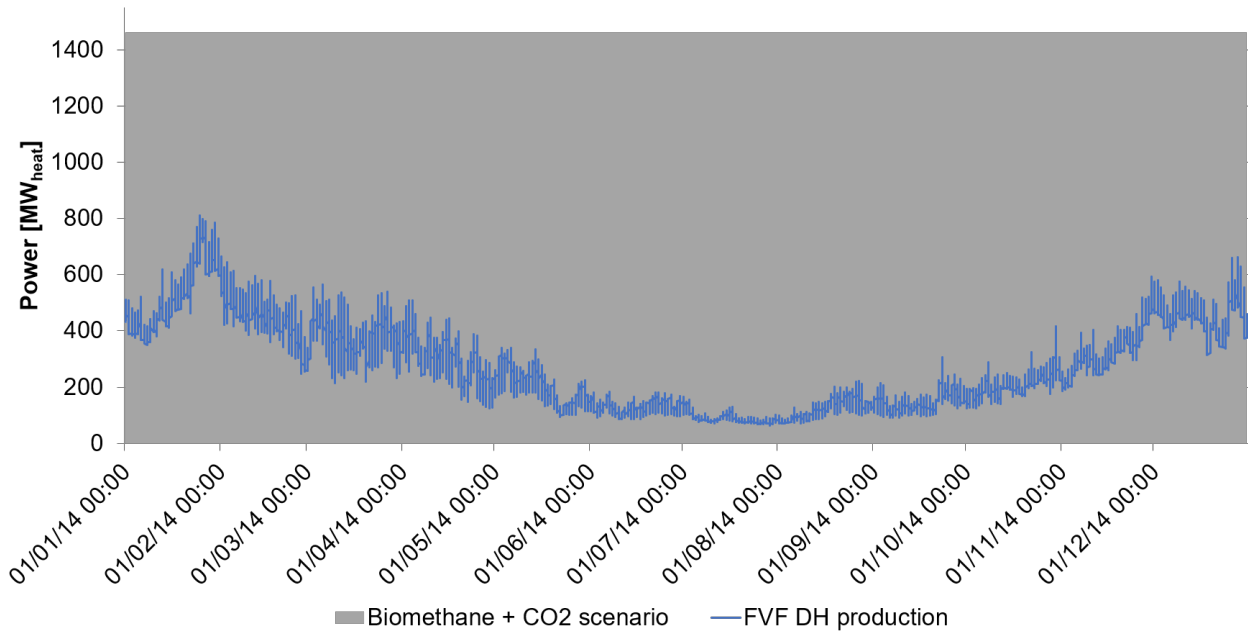


Figure 7.4: Comparison between the district heating potential in the optimal scenario with one GTL plant and the total annual district heating generation from Fjernvarme Fyn in 2014 [19]. The grey filled area represents the recoverable heat from the GTL plant when the total annual heat potential is distributed evenly over the hours of the year. The blue line is the actual heat generation from Fjernvarme Fyn.

Even in Copenhagen, the biggest city of Denmark, where the annual district heating production is around twice that of Fjernvarme Fyn in Odense, the total amount of recoverable heat cannot be used for district heating. Furthermore, due to the large size of the district heating system in Copenhagen, heat losses make it close to impossible to deliver district heating to all the consumers from one single plant [48]. With this in mind, the Odense district heating profile and size is used in the following.

In Figure 7.5, the recoverable heat from the GTL plant is scaled to be 15% of the total potential which gives an annual district heating production of 6.9 PJ. This is more realistic, but still requires an immense change to the current district heating system in Odense as it is around 80% of the production from Fjernvarme Fyn in Odense. However, in Odense, one of the last coal-fired CHP plants in Denmark is located which will be taken out of service in 2025 which makes room for new players [55].

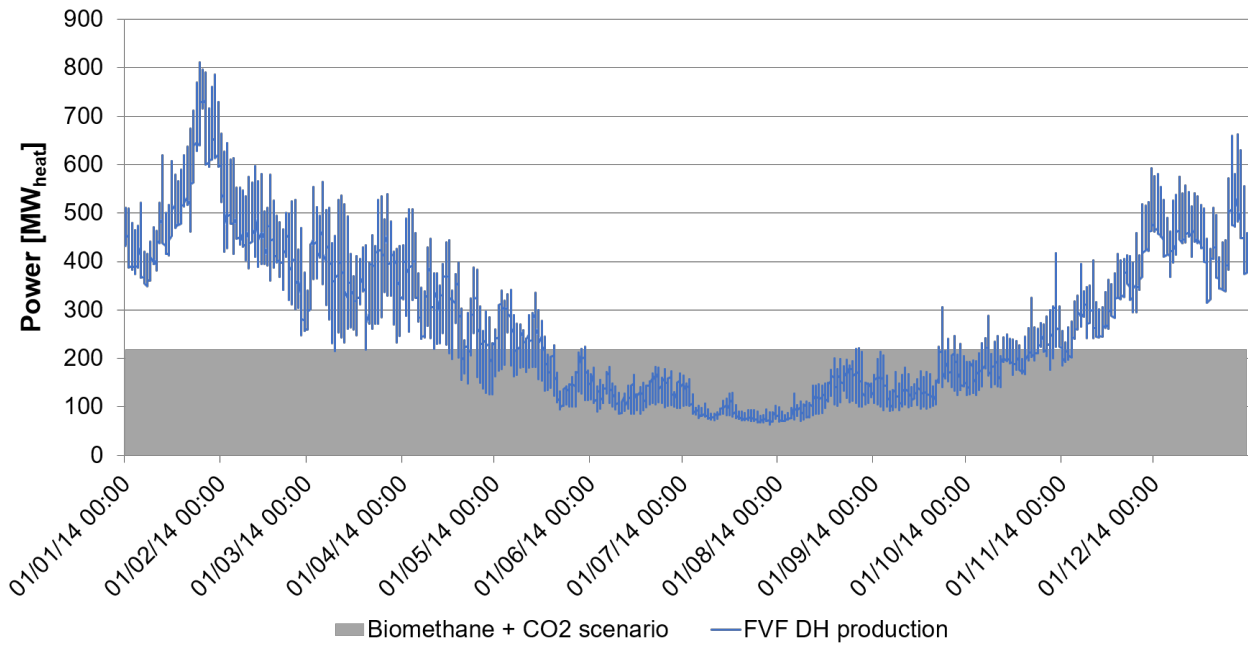


Figure 7.5: Same comparison as in Figure 7.4 but with only 15% of the district heating potential from the optimal scenario [19].

This means that only 15% of the recoverable heat will be utilised for district heating for the optimal scenario with one plant in the economic analysis. For the biomethane scenario, it will be 12%. For the biogas scenario, it will be 55%. This number is used for two main reasons. First, the annual amount of recoverable heat makes up around 55% of the total Danish base load for district heating (see Section 2.6), when this is scaled after the yearly profile of Fjernvarme Fyn. Second, one of the disadvantages of the decentralised biogas scenario is the fact that some of the GTL plants are placed away from district heating areas. For these reasons, the biogas scenario is only capable of delivering district heating equivalent to the total Danish base load.

7.2.3 Geographic locations of GTL plants in Denmark

This part discusses the location of the GTL plants in the scenarios in two parts. First, the two biomethane scenarios will be discussed together as they have the same characteristics regarding the size. Second, the biogas scenario will be discussed as this one is different from the two others regarding its feed.

The two biomethane scenarios can be made as one plant solutions with a capacity of 34,000 bpd and they require between 130.9 and 153.2 PJ of biomethane per year. The capacity of the Danish gas transmission pipes are around 317.3 PJ per year, which is more than enough to feed the

plants.¹² The large distribution pipes, which are about half the size of the transmission pipes may, therefore, also be reasonable for feeding such large plants [56]. A one plant solution is, in relation to the feed, realistic to erect somewhere along the gas transmission lines or large distribution lines. This could, for example, be in Odense, the city used above for comparison of the recoverable heat and district heating. Other large Danish cities connected to these lines are, Copenhagen, Aalborg, Esbjerg, Silkeborg, and Vejle. Most of these cities are in the southern half part of Denmark [56]. An important thing is that the city has a reasonable size so as much of the recoverable heat can be utilised for district heating.

Another point is whether the city has a coal-fired CHP plant as they are to be phased out in the coming years. As mentioned Odense has, but also Aalborg and Esbjerg have. An alternative to converting them into biomass-fired CHP plants is to install a GTL plant. Comparing the three locations, Esbjerg is worse than both Odense and Aalborg by the fact that the city is smaller than the other two. Yet another solution is to split the one plant solution into a two or three plant solution, which will result in a greater district heating income, although, it might also increase the investment costs. However, with the district heating productions for the relevant Danish cities as stated in Section 2.6, a centralised production scenario with only one or a few GTL plants is limited to Copenhagen (16.4 PJ/yr), Odense (8.6/PJ), Aalborg (6.2 PJ/yr), and maybe Esbjerg (3.4 PJ/yr) for a significant utilisation of the recoverable heat since the district heating demand in Silkeborg and Vejle are only around 1 PJ/yr. Another challenge related to the implementation of Fischer-Tropsch jet fuel production in Denmark is the many waste incineration plants. There are 26 in total with locations both in Copenhagen, Odense, Aalborg, and Esbjerg [57]. They run year round to incinerate the continuously generated waste. This may result in a lot of wasted recoverable heat from GTL plants during summer when they cover the same areas as the waste incineration plants. An option to prevent this excessive heat loss is to invest in seasonal heat storage to store the heat produced in the summer for the autumn when the demand rises again.

For the biogas scenario, the only possible solution for this to be a one plant solution is if the natural gas grid is altered to be a biogas grid instead, which is possible, yet requiring a lot of changes to the existing system where people and machines have become used to high gas quality [58]. The other solution is the solution with a lot of small GTL plants around the country close to or at the sites of the biogas plants. The result of this is a lot of decentralised GTL plants with several individual operators. The biogas plants are distributed all over Denmark, although, the concentration is bigger in Jutland and on Funen than on Zealand [54]. This distribution will be the same for the GTL plants.¹³ For this to be a viable solution the business case must be certain so most of the biogas operators will go this way, i.e. with high subsidies on the biogas for the

¹²See Appendix I for the calculation.

¹³See the distribution in Appendix J Figure J1.

GTL conversion. To ensure the proper quality of the end-products, it might be necessary to do the refining of the syncrude at centralised refineries, even though, e.g. Greyrock promises to include the distillation into final liquid products with their technology [59]. This scenario faces the same problem with the waste incineration plants as discussed above, but due to the smaller scale, some plants can also be located in district heating areas without waste incineration. At the same time, it can also be discussed if it must be placed so far out the countryside that none of the recoverable heat can be utilised for district heating. Still the most uncertain point about this scenario is that the investment decision will be spread out on a lot of people, and not just one or two investors, which makes it harder for politicians to predict how fast it will happen, in comparison to a scenario where they have just a few stakeholders to work with.

From this part of the analysis, it can be concluded that the optimal scenario also has some advantages over the others in relation to the implementation of the technology in the Danish system. This is due to its better performance and the fact that it requires one or only a few operators. Its merits increase further if it is split into a few plants so the discussed 15% utilisation of the recoverable heat can be increased. Despite this, it is the one plant solution that is evaluated next in the economic analysis.

7.3 Economic analysis

The economic analysis is made on the basis of the performance data from the Aspen Plus simulations and with the above numbers regarding the district heating. The CAPEX used for the biomethane and biomethane and CO₂ scenarios are equivalent to the CAPEX described in Section 2.7.1 because a one plant solution will be approximately the same size as the reference. For the biogas scenario, the distributed production sites will increase the CAPEX, so they will be 1.5 times more expensive on the specific investment as stated in Section 2.7. The feed costs are as described in Section 2.7.2 without subsidies. Afterwards, the size of the subsidies will be discussed and related to the present-day figures. The economic analysis is comprised of the LCOE, a feed to product cost relationship, and the business and socioeconomic NPVs.

7.3.1 Levelised cost of energy

Figure 7.6 shows the price composition of the LCOE of the liquid products.¹⁴ The most significant cost driver for the GTL conversion is the feed, but also the CO₂ cost in the third scenario is significantly larger than the other costs. Notice that the most expensive scenario is the one with the optimal feed ratio, although, under the premise of the high carbon capture cost. If this is

¹⁴See Appendix G for GTL calculation of the different costs.

halved as predicted [24], the performance will be equivalent to biomethane scenario cost wise, but better technology wise.

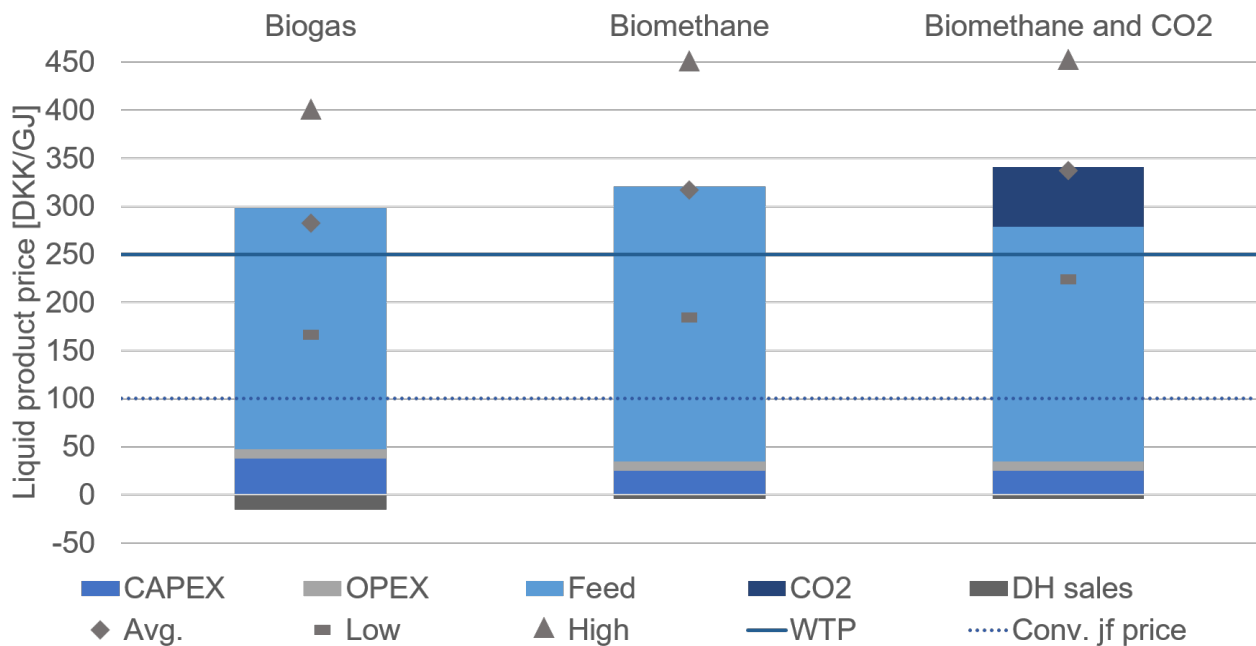


Figure 7.6: Price composition of GTL liquid products where the band between *high* and *low* shows the price interval resulting from the sensitivity on the methane input price. Feed refers to the cost of biogas or biomethane. The final price for bio jet fuel is only valid if it is the same for biogasoline. The *Avg.* price is based on a feed cost of 130 DKK/GJ, the *Low* price is 70 DKK/GJ, and the *High* price is 190 DKK/GJ.

For all of the scenarios, the lowest half of the price interval ranges from 70-130 DKK/GJ input, whereas the top half of it ranges from 130-190 DKK/GJ input. Recall the typical biogas production costs are between 130-142 DKK/GJ and the cost for methanation between 162-189 DKK/GJ. So, for the biogas scenario, about the lowest half of the price interval requires subsidies whereas it for the biomethane scenarios are closer to the lowest three-quarters. This means that it is only considered realistic to get below the WTP line at 250 DKK/GJ with subsidies. However, this WTP of 250 DKK/GJ originates from a statement given in relation to a demonstration plant, so it can also be argued that the WTP for the total Danish jet fuel demand will be somewhat lower [25]. Evidently, the lower it gets, the more reliant the scenarios are on subsidies. At the same time, the graph assumes that all of the liquid products from the GTL plant has the same price. With the energy tax on *gasoline equivalents*, the main WTP increase on the biogasoline will probably be due to its image value. For the personal transport fleet, this may not be significant enough to make a difference. Be it said, the optimal use for the biogasoline in the future may not be the personal transport as this is easier to electrify, which means other sectors in need of biofuel may be willing to pay the necessary premium price on the biogasoline. These sectors are discussed further down in this section.

The blue vertical dotted line in Figure 7.6 is the conventional jet fuel price of 100 DKK/GJ. It is obvious that this price is not high enough to make any of the scenarios feasible, which means that the WTP for biofuels must be significantly higher than this for this type of biofuels to be a profitable business, even if it is subsidised. A Swedish study from 2018 [60] showed that Swedish company organisations are only willing to pay a premium of 11.9% on top of the conventional ticket on average if a bio/conventional jet fuel blend of 50/50 was used [60]. With fuel cost constituting up to 50% of the operating cost of an airline, an increase in the jet fuel price will have a significant impact on the ticket prices passed on to the passengers [61]. These findings are interesting and show the difficulties the aviation industry is facing if it wants to turn green. At the same time, they also show that if the price of jet fuel is doubled, the flight tickets will only change by around 25%, which is only around twice the average increase in WTP found from the Swedish study — and notice that the investigated jet fuel in the current study is 100% bio jet fuel.

7.3.2 Relationship between feed cost and liquid product price

The huge impact the feed cost has on the LCOE makes it interesting to examine the relationship between the feed cost and the liquid product price. With all other cost parameters held constant, the relationship can be seen in Figure 7.7 where the three scenarios are depicted together. The lines in the figure can be used to get a better understanding of the liquid product price that can be produced by GTL plants at different points in time. With subsidies in the order of the current in the sector, see Section 2.7.2, a possible feed price to a GTL plant could be 90 DKK/GJ.

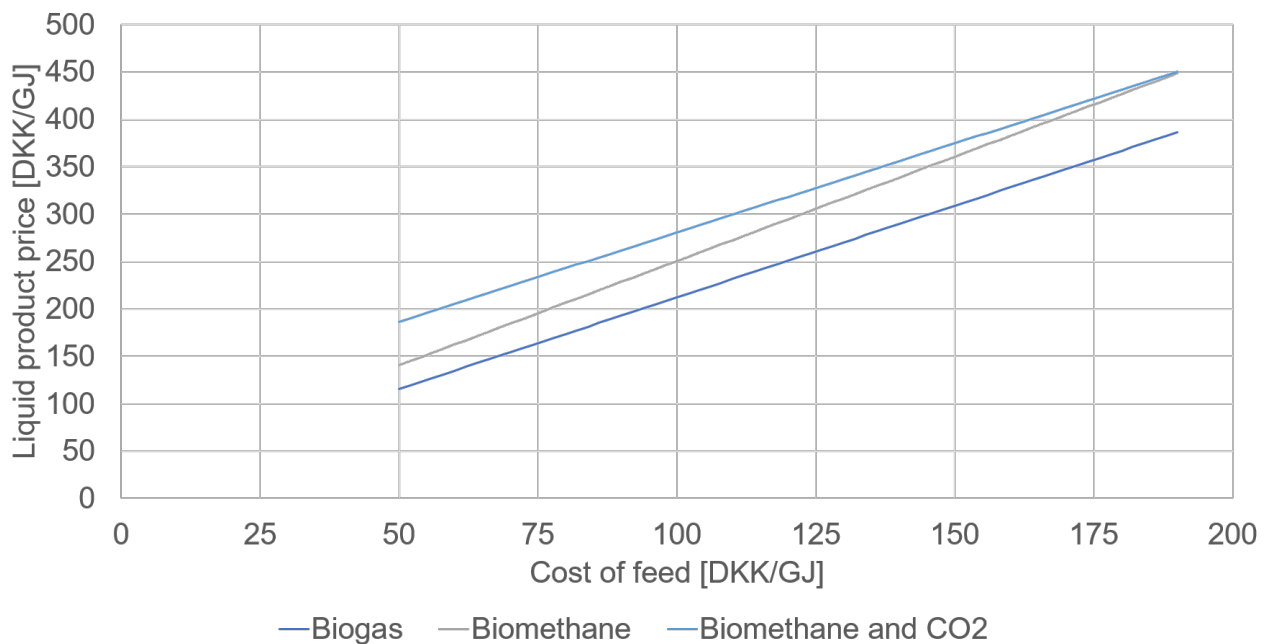


Figure 7.7: Relationship between feed cost and liquid product price. The x-axis is the cost of the feed, and y-axis is the price of the liquid products from the GTL plant in the three scenarios.

In the graph and with a feed cost of 90 DKK/GJ, the resulting liquid product price is between 180-260 DKK/GJ. On the other hand, a worst-case scenario could imply the highest possible technology costs (e.g. for methanation) and no subsidies. This gives a feed cost of 190 DKK/GJ resulting in a liquid product price ranging from 375-450 DKK/GJ. By looking into the energy agreement at the time of investment, it is, however, possible to get a greater certainty on the points regarding the subsidies, so this worst case scenario can be avoided. The relationship in Figure 7.7 can also be used by an investor to see the possible liquid product price they will be able to deliver if they can be guaranteed a specified feed cost to their GTL plant.

7.3.3 Net present value

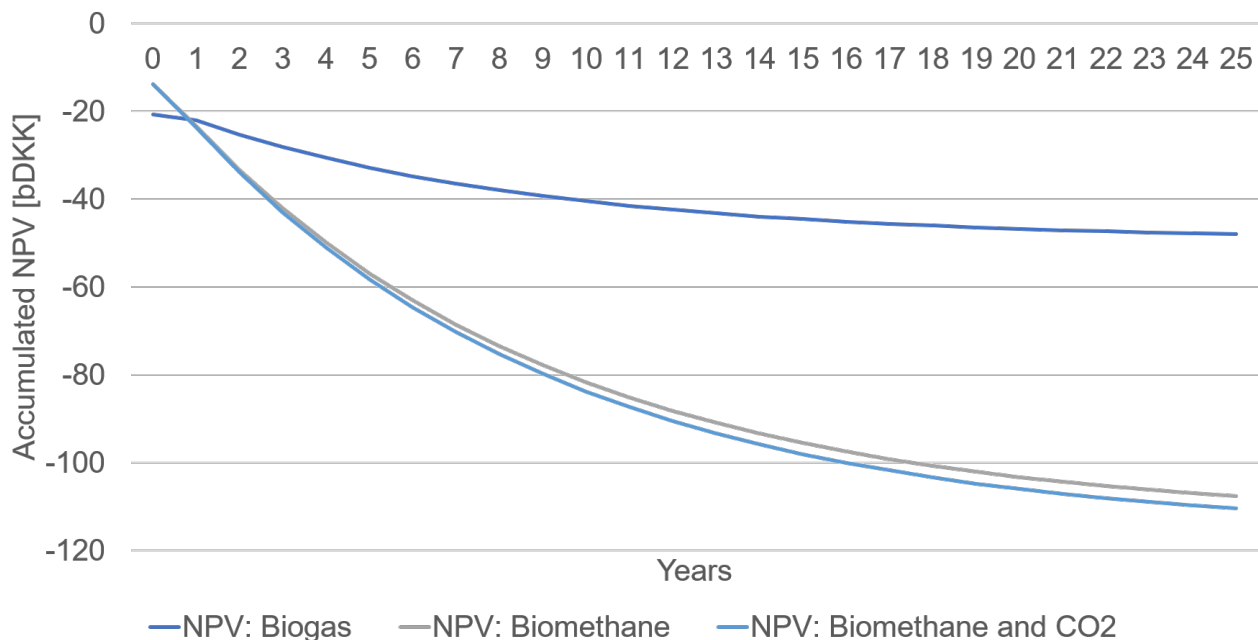


Figure 7.8: Business economic NPV over 25 years for the three scenarios.

In Figure 7.8, the NPV over 25 years is depicted without subsidies rewarded on neither biogas nor biomethane. Without subsidies, none of the scenarios yields a positive NPV after 25 years. In addition, none of them does not even have an annual revenue which can be seen from the downwards slopes of the NPV graphs. This makes investments in the Fischer-Tropsch jet fuel technology highly unattractable when subsidies are disregarded. For the three scenarios, the subsidies on the feed should be 48.43 DKK/GJ, 98.80 DK/GJ, and 118.78 DKK/GJ to have an NPV after 25 of zero for the biogas, biomethane, and biomethane and CO₂ scenarios, respectively. However, comparing the present-day subsidies with these figures, it is not unrealistic to achieve a positive business case for Fischer-Tropsch jet fuel production, although, some changes must be

made since the subsidies as of today are not rewarded completely in the scenarios. In Appendix H, the NPVs of the three scenarios are depicted with their respective breakeven subsidy.

7.3.4 Socioeconomic analysis

The socioeconomic NPVs resembles the NPVs in the business economics where the subsidies are removed (as in Figure 7.8), i.e. all of the scenarios have negative NPVs. The socioeconomic NPVs are calculated based on the numbers given in Section 2.7 and methods in Section 3.2.

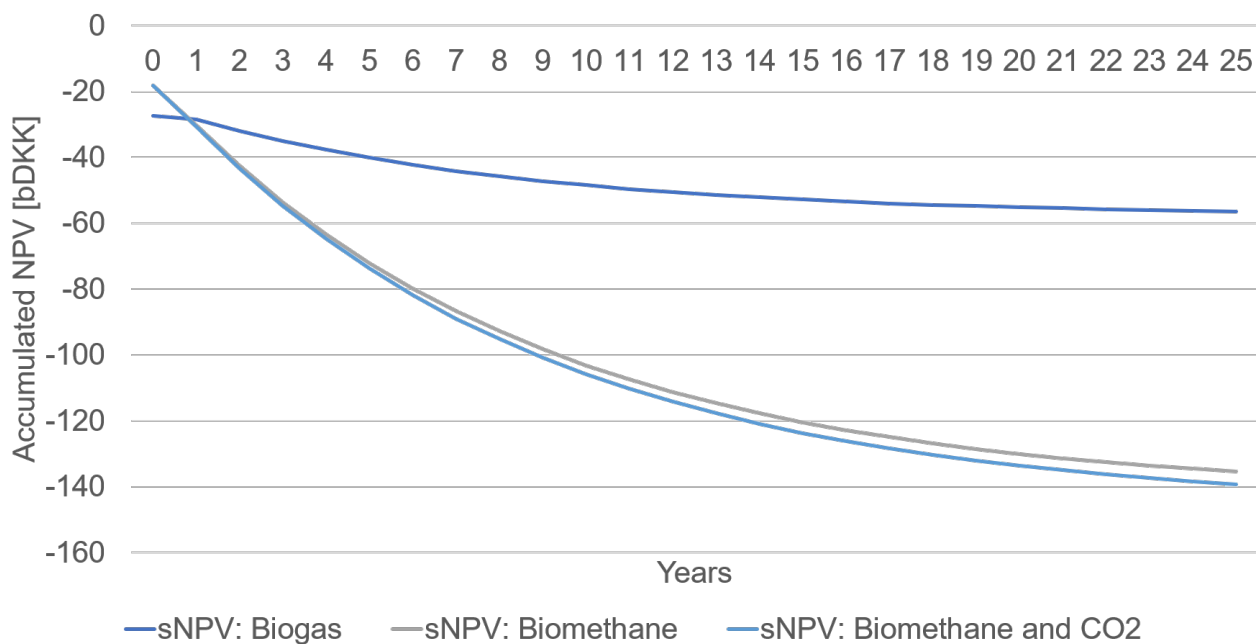


Figure 7.9: Socioeconomic NPV over 25 years for the three scenarios.

Figure 7.9 shows the development of the socioeconomic NPVs for the three scenarios over a lifetime of 25 years. The socioeconomic NPV calculations show that if the harm traditional jet fuel imposes on society is equivalent to the CO₂ quota price, it does not make it up for the additional cost of producing Fischer-Tropsch jet fuels no matter which of the scenarios are followed. For the socioeconomic NPVs to be positive, the harm costs from CO₂ should be 8.9, 20.0, and 20.6 times greater than 200 DKK/ton (i.e. 1780 DKK/ton, 4000 DKK/ton, and 4120 DKK/ton) for the biogas, biomethane and biomethane and CO₂ scenarios, respectively. In relation to this, it can be discussed if the CO₂ quota price does show the harm that CO₂ emission impose on society, and if each ton of CO₂ causes the same damage as the previous.

Despite this, with the goal to become fossil free at latest by 2050, a fossil scenario is not an alternative, so the results from this analysis can be used to compare this pathway of making the aviation industry fossil free with other pathways to find the solution with the lowest CO₂ abatement cost.

From the above, the economically best solution is the biogas scenario. This means, that there is a discrepancy between the technically and economically best scenarios for Fischer-Tropsch jet fuel production in Denmark. In the end, the best solution, may therefore, be the biomethane scenario due to the fact that it shares some of the advantages of the biomethane with CO₂ scenario, but at a lower cost, although, only slight improvements of the CO₂ capture technology is required to make these two equal.

8 Conclusion

Implementation of Fischer-Tropsch jet fuel production in the Danish energy system is not unrealistic per se. However, when implementing it in such a large scale as to satisfy the entire demand of the Danish aviation industry, difficulties appear in several forms. First, the biogas availability is too low today. Moreover, the theoretical potential is only sufficient if all of it is used to produce Fischer-Tropsch jet fuel. At the same time, other sectors are also using biogas or biomethane in a zero-emission society. This requires careful and strategic energy planning if this proposal is implemented in full scale. Alternatives to satisfy the jet fuel demand include; the addition of syngas made from co-electrolysis, methanation of captured CO₂, and import of a share of the biokerosene needed to satisfy the demand.

The potential in the recoverable heat from the Fischer-Tropsch jet fuel production is significant, although, too large to be fully utilised in the Danish energy system. In the analysed scenarios, the utilisation of the recoverable heat ranges from 12% to 55% where the location and number of GTL plants are the main determinants. The lowest utilisation was found to be for one large central GTL plant even when located in a reasonable district heating area. A multiple plant solution will, therefore, yield a better utilisation of the recoverable heat.

Economically, the production of Fischer-Tropsch jet fuel relies on subsidies. For a scenario where the feed to the GTL conversion is biogas, the required subsidy for the NPV to be zero is 48.43 DKK/GJ. When the feed is biomethane it is 98.80 DKK/GJ, and when it is biomethane with added CO₂ it must be 118.78 DKK/GJ. With the current prices for carbon capture, the optimisation of the process by CO₂ co-feed does not pay off, but it may be competitive if carbon capture costs are halved. The least expensive scenario is a scenario with many distributed small-scale GTL plants using biogas despite higher investment costs. From a socioeconomic point of view, the implementation of Fischer-Tropsch jet fuel in Denmark is a negative initiative with the CO₂ quota price of 200 DKK/ton as the social harm cost. If the harm, however, is 1780 DKK/ton, 4000 DKK/ton, and 4120 DKK/ton, the three scenarios get positive values from socioeconomic point of view.

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Appendices

A Aspen Plus configuration

The thermodynamic property model NRTL-HOC (non-random two-liquid model with Hayden O'Connell vapor fugacity coefficient) is used for the Aspen Plus model [62]. R-GIBBS reactors in Aspen Plus use Gibbs energy minimisation for the reactions. Theory about Gibbs energy minimisation can be found from *Y. A. Çengel and M. A. Boles* [44].

Table A1: List of available products and reactants in the Aspen Plus model.

Number	Component name
1	HYDROGEN
2	OXYGEN
3	NITROGEN
4	WATER
5	CO
6	CO2
7	METHANE
8	UNDECANE
9	OCTANE
10	PROPANE
11	BUTANE

Table A2: Overview if the configuration of the main blocks in Aspen Plus. Possible products refers to table A1. The base method used in the simulation is NRTL-HOC [62].

Reformer (R-GIBBS) [7]	
Temperature [°C]	900
Pressure [bar]	20
Possible products	[1:11]
Fischer-Tropsch reactor [5, ch. 4]:	
Jet fuel (R-GIBBS)	
Temperature [°C]	200
Pressure [bar]	10
Possible products	[1,4,6,8]
Motor gasoline (R-GIBBS)	
Temperature [°C]	200
Pressure [bar]	10
Possible products	[1,4,6,9]
Other products (R-GIBBS)	
Temperature [°C]	200
Pressure [bar]	10
Possible products	[1,4,6,10,11]
Syngas splitter (F-SPLIT)	
Jet fuel (fraction)	59%
Motor gasoline (fraction)	35%
Other products (fraction)	6%
Heat exchangers for DH (gas-to-liquid)	
Hot/cold temperature difference [K]	15
DH hot temperature [°C]	70
DH pressure [bar]	5
Heat exchangers for DH (liquid-to-liquid)	
Hot/cold temperature difference [K]	10
DH hot temperature [°C]	70
DH pressure [bar]	5
Reformer heater (R-GIBBS)	
Temperature [°C]	1150
Pressure [bar]	1
Air (stream)	
Nitrogen (fraction)	79%
Oxygen (fraction)	21%

B Calculation of optimal H₂:CO ratio

How the molar fraction in the syncrude is calculated based on the molar masses and mass fraction in the syncrude.

$$\begin{aligned}\text{moles/ton}_{\text{part}} &= \frac{\text{mass}\%}{M} \\ \text{mole}\% &= \frac{\text{moles/ton}_{\text{part}}}{\text{moles/ton}_{\text{total}}}\end{aligned}$$

This gives:

$$\begin{aligned}\text{C}_{11}\text{H}_{24} : & \frac{59\%}{156 \text{ ton/Mmol}} = 3782 \text{ moles/ton} \\ \text{C}_8\text{H}_{18} : & \frac{35\%}{114 \text{ ton/Mmol}} = 3070 \text{ moles/ton} \\ \text{C}_{11}\text{H}_{24} : & \frac{3782 \text{ moles/ton}}{6852 \text{ moles/ton}} = 55\% \\ \text{C}_8\text{H}_{18} : & \frac{3070 \text{ moles/ton}}{6852 \text{ moles/ton}} = 45\%\end{aligned}$$

Calculation of optimal ratio of the evaluated syncrude with the above composition: As we are only working with alkanes with the chemical formula C_nH_{n+2} and as we want to convert all of the carbon in the CO into useful products (in theory), the number of H₂ should be two times the amount CO plus one. This is to remove the amount of excess H₂. This means that all the hydrogen that does not go into the hydrocarbons reacts with the oxygen from the CO and makes water. This gives:

$$\begin{aligned}\text{C}_{11}\text{H}_{24} : \quad n_{\text{H}_2} &= 2 \cdot 11 + 1 = 23\text{H}_2 \\ \text{C}_8\text{H}_{18} : \quad n_{\text{H}_2} &= 2 \cdot 8 + 1 = 17\text{H}_2\end{aligned}$$

When the optimal H₂:CO ratio is calculated and weighted we get:

$$\begin{aligned}\text{C}_{11}\text{H}_{24} : \quad \frac{23\text{H}_2}{11\text{CO}} &= 2.0909 \\ \text{C}_8\text{H}_{18} : \quad \frac{17\text{H}_2}{8\text{CO}} &= 2.125 \\ \text{weighted :} \quad &55\% \cdot 2.0909 + 45\% \cdot 2.125 = 2.1062\end{aligned}$$

So the optimal ratio is 2.1062 when the simplification current simplification is made.

C Pre-study supplements

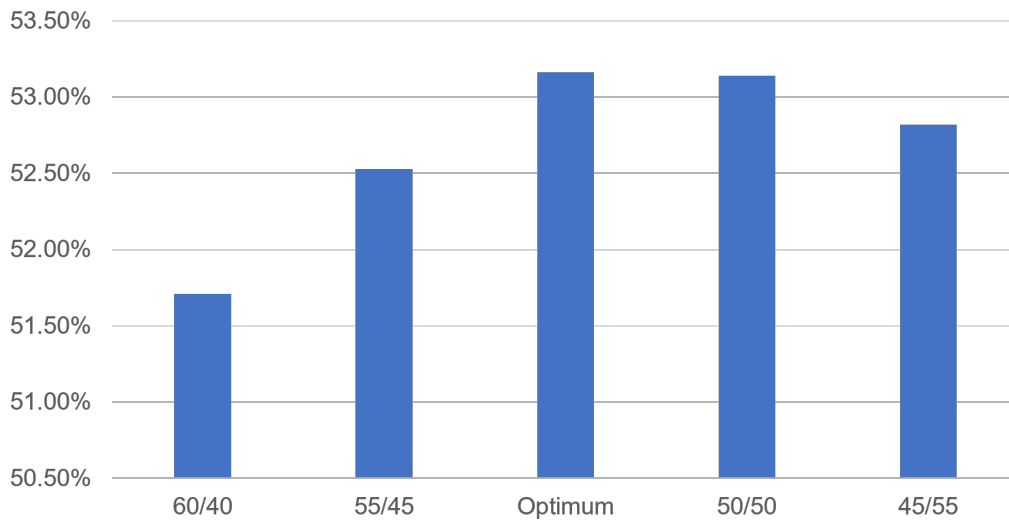


Figure C1: Product efficiency for different setups with varying CH₄:CO₂ ratio. The optimal feed ratio is found from *user defined* blocks and has the label *Optimal* which is 51:49 CH₄:CO₂.

C.1 Electrical work in the GTL plant

SET UP: A pump is inserted to compress the water to be equal to the pressure in the reformer (20 bar). The pressure at the syngas separation unit is lowered to just above atmospheric pressure (1.2 bar) to simulate a PSA unit and it is then raised again to the pressure in the Fischer-Tropsch reactor (10 bar).

The differences between the model with and without the pump work included are:

- A pump before the reformer raising the pressure of the water from 1 bar to 20 bar.
- A valve in relation to the syngas conditioning at 1.2 bar (slightly above atmospheric pressure).
- Additional pump work before the Fischer-Tropsch synthesis section to increase it to 10 bar.
- The energy from the separation unit is included in the pump work model and not in the other one.

Comparing the results with and without pumps, compressors, and separation shows that the change of the efficiency is minor and they do not constitute a significant energy consumption in the overall GTL process. See the comparison in Figure C2. This justifies the simplification of neglecting the energy consumption of the pumps and compressors in the system.

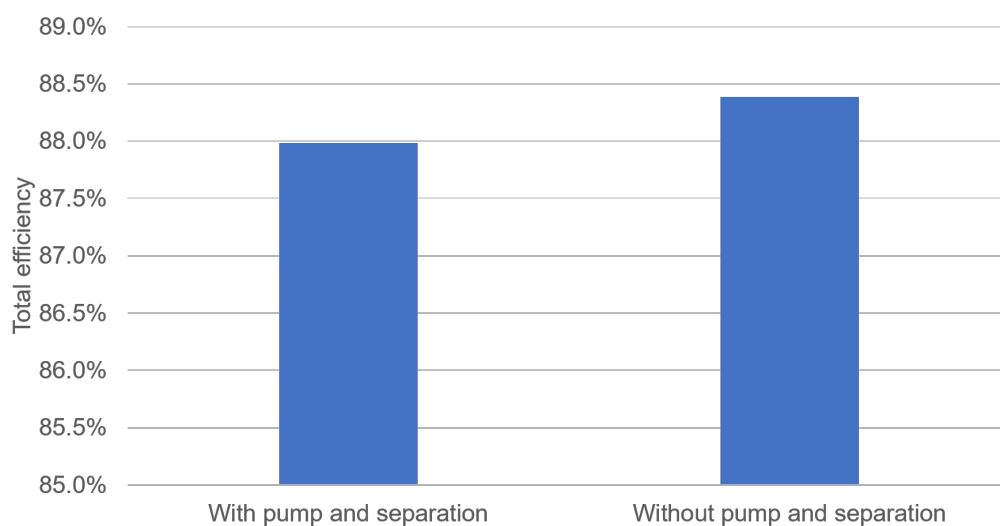


Figure C2: Effects on efficiency of pumps, compressors, and separation in the Aspen Plus models. Left column shows the efficiency with, and right without.

D Plant size calculation

The size of the plant is calculated on the assumption that it should run constantly throughout the whole year. One bbl is equivalent to 5.4 GJ [63]. The total liquid energy production when the total aviation industry should have the jet fuel demand covered is 69.6 PJ per year. This gives

$$\text{plant size} = \frac{69.6 \text{ PJ/yr}}{365 \text{ days/yr}} \cdot \frac{10^6 \text{ GJ/PJ}}{5.4 \text{ GJ/bbl}} = 33,790 \text{ bbl/day} \quad (8)$$

So for a one plant solution the capacity of the plant should be 33,790 bbl/day to meet the Danish demand for bio jet fuel.

E District heating calculations

E.1 Base load of district heating areas

To calculate the base load for district heating areas where the only available data is the total annual heat production, the total annual heat production divided by the minimum heat delivery during the year for Fjernvarme Fyn [19]. The total production in 2014 was 2,560,786 MWh, the minimum hour heat production was 58.91 MWh.

$$R = \frac{2,560,786 \text{ MWh}}{58.91 \text{ MWh/h}} = 43,466.96 \text{ h} \quad (9)$$

With this ratio the base load of a given city can be calculated by dividing the total yearly heat production in MWh by R .

E.2 Heat production from scenarios

The hourly heat production from the scenarios are calculated as

$$P_{heat} [\text{MW}] = \frac{E_{year} [\text{PJ/yr}]}{3.6 \cdot 10^{-6} [\text{PJ/MWh}] \cdot 8760 [\text{h/yr}]} \quad (10)$$

F District heating base load in Denmark

In 2017 the total Danish district heating sector produced 135 PJ [15]. To find the annual base load, the value must be converted to MWh to be able to use relation for the base load found from Fjernvarme Fyn in Appendix E.1.

$$\text{Ann. DK tot. [MWh]} = \frac{135 \text{ PJ}}{3.6 \cdot 10^{-6} \text{ PJ/MWh}} = 37,500,000 \text{ MWh} \quad (11)$$

The base load is

$$\text{Base load [MW]} = \frac{37,500,000 \text{ MWh}}{43,466.96 \text{ h}} = 862.7 \text{ MW} \quad (12)$$

The annual base load is then

$$\text{Ann. base load [MWh]} = 862.7 \text{ MW} \cdot 8760 \text{ h} = 7,557,464.34 \text{ MWh} \quad (13)$$

Converting back to a value in PJ, the annual base load is

$$\text{Ann. base load [PJ]} = 7,557,464.34 \text{ MWh} \cdot 3.6 \cdot 10^{-6} \text{ PJ/MWh} = 27.2069 \quad (14)$$

So, the annual base load in the entire country of Denmark is 27.21 PJ when it is scaled after the profile of Fjernvarme Fyn in Odense.

G Jet fuel price calculation

CAPEX:

The jet fuel price composition is calculated based on investment cost of 62,000 USD/bpd (2010) [5].

With 5.4 GJ/bbl this gives $\frac{62,000 \text{ USD/bpd}}{5.4 \text{ GJ/bpd}} = 11481 \text{ USD/GJ}$.

Converted into DKK by the avg. exchange rate in 2010, 5.6186 DKK/USD [40], this is 64,510 DKK/GJ/day equivalent to 177 DKK/GJ/year.

With the net price index the 2019/2010 factor is 1.117647 [42] giving 198 DKK/GJ/day (2019).

Amortised over 25 years at 12% IR it is 25 DKK/GJ/year.

OPEX:

OPEX is 8.7 USD/GJ (2011).

Average exchange rate in 2011 was 5.3552 DKK/USD [41].

2019/2011 ratio is 1.090329. This gives

$$\text{OPEX} = \frac{8.7 \text{ USD}_{2011}/\text{GJ} \cdot 5.3552 \text{ DKK}_{2011}/\text{USD}_{2011} \cdot 1.090329 \text{ yr}_{2019}/\text{yr}_{2011}}{5.4 \text{ GJ/bbl}} \quad (15)$$

Fuel cost:

Ranging from 70-190 DKK/GJ.

As the above prices are for the output this needs to be converted to be relatable to the output as well. Done by division by the efficiencies for the respective scenarios: biogas = 52%, biomethane = 45%, and biomethane and CO₂ = 53%.

CO₂ cost:

The used price for CO₂ is 600 DKK/ton.

The weight of methane is the $LHV^{-1} = [50 \text{ MJ/kg}]^{-1} = 20 \text{ kg/GJ}$.

The molar mass ratio $\frac{CO_2}{CH_4} = \frac{44}{16} = 2.8$.

The price at a 50/50 blend is then,

$$CO_2 \text{ cost} = 0.6 \text{ DKK/kg} \cdot 20 \text{ kg/GJ}_{CH_4} \cdot 2.8 M_{CO_2}/M_{CH_4} = 33 \text{ DKK/GJ}_{CH_4} \quad (16)$$

This is divided by the efficiency of the biomethane and CO2 scenario, which yields

$$CO_{2 \text{ cost}} = \frac{33 \text{ DKK/GJ}_{CH_4}}{53\%} = 62.1 \text{ DKK/GJ}_{output} \quad (17)$$

District heating sales:

From the models the DH heating output per liquid product output (district heating efficiency to product efficiency ratio) was found to be

$$0.70 \text{ GJ}_{DH_{out}}/\text{GJ}_{output} \text{ (biogas)}$$

$$0.86 \text{ GJ}_{DH_{out}}/\text{GJ}_{output} \text{ (biomethane)}$$

$$0.66 \text{ GJ}_{DH_{out}}/\text{GJ}_{output} \text{ (biomethane and CO2)}$$

The average remuneration is 141 DKK/MWh equal to 40.7 DKK/GJ [33]. The district heating income is then calculated as:

$$DH_{income, biogas} = \frac{40.7 \text{ DKK/GJ}_{DH}}{0.70 \text{ GJ}_{DH}/\text{GJ}_{out}} \cdot 55\% = 29 \text{ DKK/GJ}_{out} \quad (18)$$

$$DH_{income, biomethane} = \frac{40.7 \text{ DKK/GJ}_{DH}}{0.86 \text{ GJ}_{DH}/\text{GJ}_{out}} \cdot 12\% = 4 \text{ DKK/GJ}_{out} \quad (19)$$

$$DH_{income, biomethane+CO_2} = \frac{40.7 \text{ DKK/GJ}_{DH}}{0.66 \text{ GJ}_{DH}/\text{GJ}_{out}} \cdot 15\% = 4 \text{ DKK/GJ}_{out} \quad (20)$$

Where the percentages refer to the discussion in the report (section 7.2.2).

Converting into MW_{input} :

The above prices per GJ_{output} can be converted into MWh_{input} by multiplying by 3.6, since 3.6 GJ is 1 MWh. For the CAPEX the cost should be further multiplied by 8760 h/yr, to give the CAPEX per installed capacity in MW.

H Subsidies required to give a positive NPV for the three scenarios

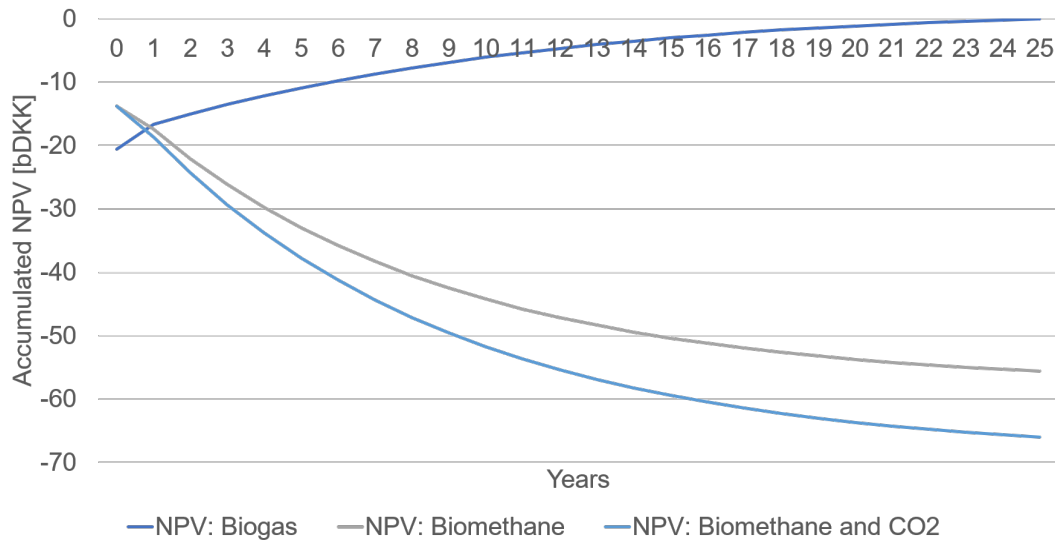


Figure H1: Business economic NPV with a subsidy of 48.42 DKK/GJ on the feed making the biogas scenario feasible.

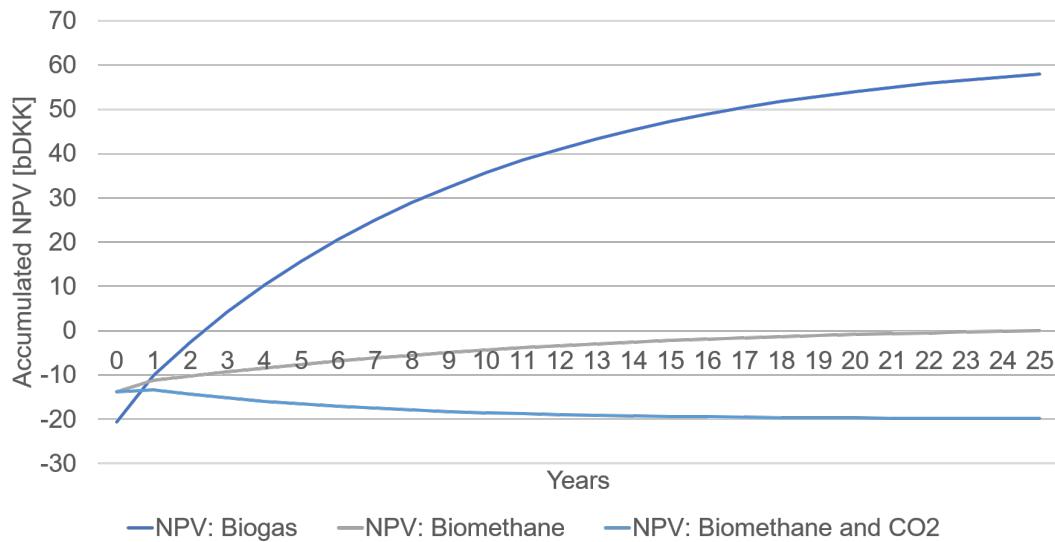


Figure H2: Business economic NPV with a subsidy of 98.79 DKK/GJ on the feed making the biogas and biomethane scenarios feasible.

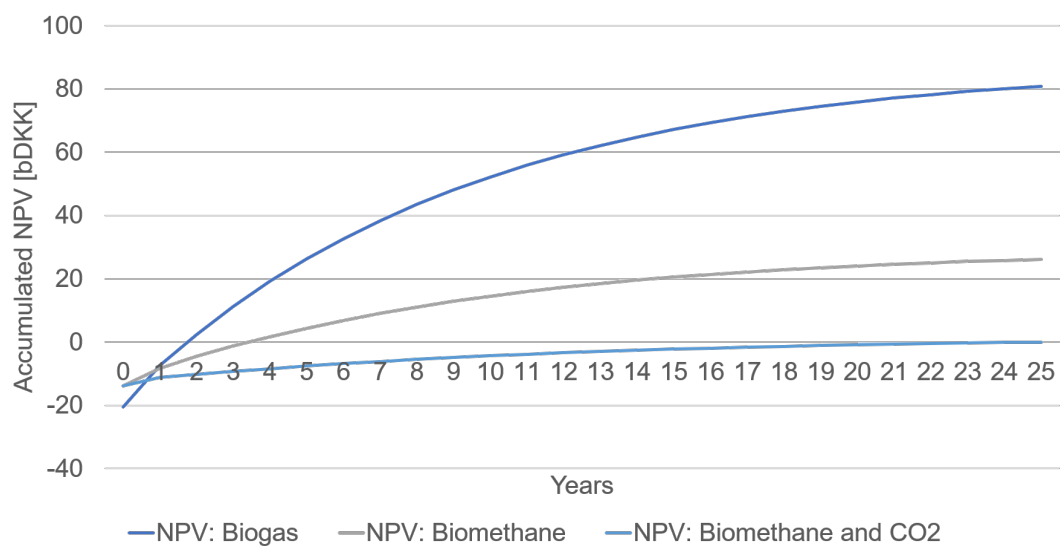


Figure H3: Business economic NPV with a subsidy of 118.78 DKK/GJ on the feed making all three scenarios feasible.

I Gas system in Denmark

The capacity of the Danish gas transmission system is 8 billion Nm^3 per year. With an LHV of $39.66 \text{ MJ}/\text{Nm}^3$ the annual energy capacity of the system can be calculated to

$$8 \text{ bNm}^3 \cdot 39.66 \text{ MJ}/\text{Nm}^3 \cdot 10^{-9} \text{ PJ}/\text{GJ} = 317.28 \text{ PJ} \quad (21)$$

So each year 317.28 PJ can be transmitted through the natural gas grid transmission system. For the large distribution pipes, it is reasonable to assumed it is around half of this [56].

J Biogas plants in Denmark

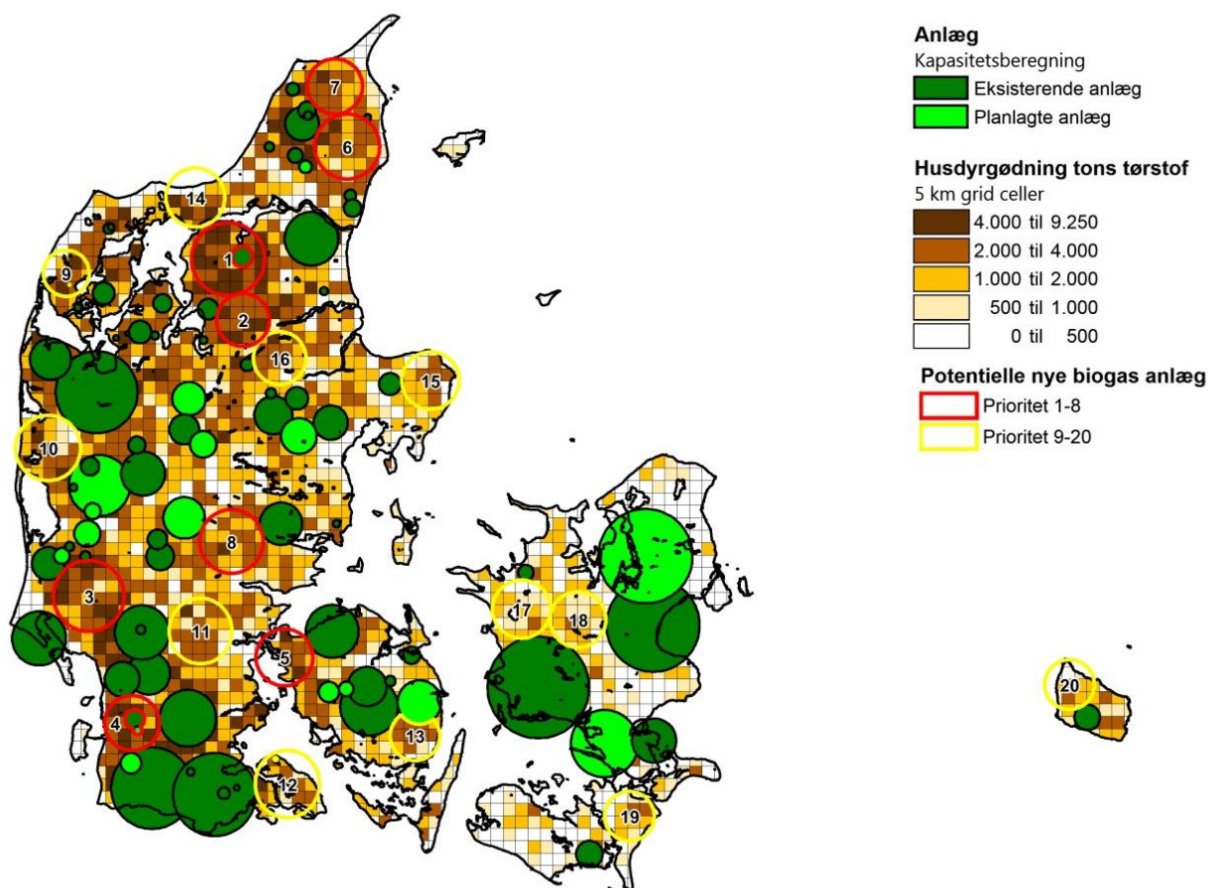


Figure J1: Present and planned biogas plants in Denmark. The figure is taken from the report *Kortlægning af hensigtsmæssig lokalisering af nye biogasanlæg i Danmark*, p. 5 [54, p. 5].

K Heat streams before and after heat exchange

Figure K1 show the hot and cold streams in the Aspen Plus model of the reformer and the Fischer-Tropsch synthesis reactor before heat exchange.

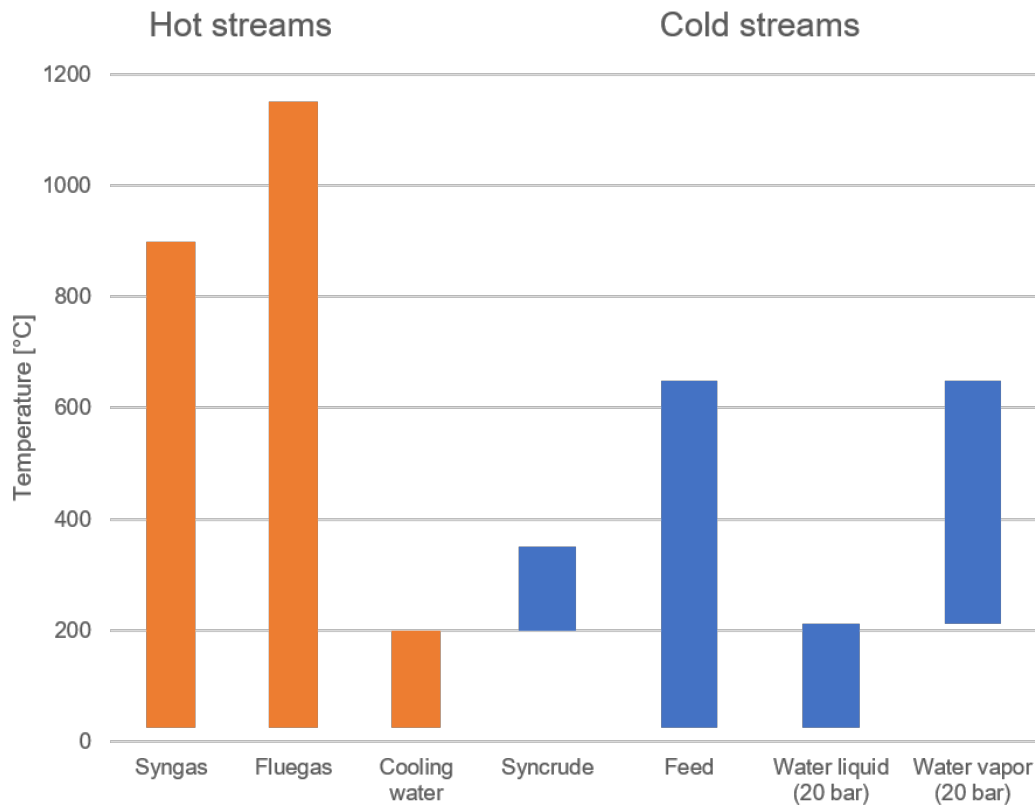


Figure K1: Hot at cold streams before heat exchange.

Common for all of the three scenarios is that the heat recuperation help eliminate the need for any additional pre-heating. The temperatures of the hot streams after the internal heat exchange vary where they are lowest in the optimal biomethane and CO₂ scenario and highest in the biomethane scenario. The heat streams after heat recuperation are exemplified in Figure K2 for the biogas scenario where they are as shown in Figure K1 before. That is, all of the heat streams posses a potential for delivering district heating. The same goes for the other two scenarios.

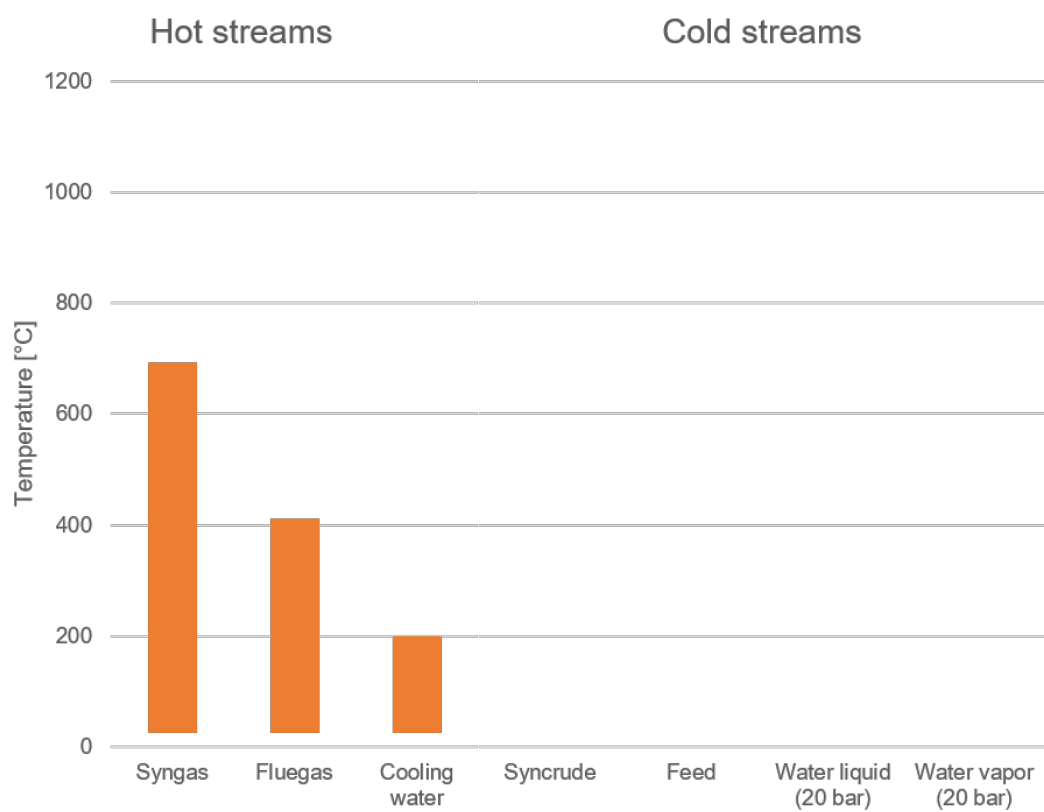


Figure K2: Hot and cold streams after heat exchange and before district heating.

L Stoichiometric calculations vs. Aspen Plus simulation

Properties

Thermodynamic and chemical properties

	Molar mass [g/mol]	LHV [MJ/kg]	LHV [kJ/mol]
H2	2	120.0	239.9
O2	16	0.0	0.0
CO	28	10.1	282.8
CO2	44	0.0	0.0
H2O (g)	18	0.0	0.0
CH4	16	50.0	800.1
C11H24	156.3	44.2	6907.5

Ratio 1:0

Steam methane reforming

Reaction: 11 CH₄ + 11 H₂O ---> 11 CO + 33 H₂

Energy [kJ] 8802 + 0 ---> 3111 + 7917

Heat out [kJ] -2227

AP Heat duty [Watt] -619

AP Energy [kJ] -2228

Rel. difference 0.1%

Jet Fuel production, Fischer-Tropsch:

Reaction: 11 CO + 33 H₂ ---> 1 C₁₁H₂₄ + 11 H₂O + 10 H₂

Energy [kJ] 3111 + 7917 6908 0 2399

Heat out [kJ] 1721

AP Heat duty [Watt] 479

AP Energy [kJ] 1724

Rel. difference 0.2%

Ratio 3:1

Carbon input: **11** 0.772725 H2:CO
2.1

Steam methane reforming

Reaction:	8.499975 CH4	+	2.5 CO2	+	6.0 H2O	--->	11 CO	+	23.0 H2
Energy [kJ]	6801	+	0	+	0	--->	3111	+	5518

Heat out [kJ] **-1828**
AP Heat duty [Watt] -509
AP Energy [kJ] -1833
Rel. difference 0.3%

Jet Fuel production, Fischer-Tropsch:

Reaction:	11 CO	+	23 H2	--->	1.0 C11H24	+	11 H2O	+	0 H2	+	0.0 CO2
Energy [kJ]	3111	+	5518		6908		0		0		0

Heat out [kJ] **1721**
AP Heat duty [Watt] 481
AP Energy [kJ] 1732
Rel. difference 0.6%

Ratio 3:2

Carbon input: 15

Steam methane reforming

Reaction:	9 CH ₄	+	6 CO ₂	+	3 H ₂ O	--->	15 CO	+	21 H ₂
Energy [kJ]	7201	+	0	+	0	--->	4242	+	5038

Heat out [kJ] -2079

AP Heat duty [Watt] -581

AP Energy [kJ] -2092

Rel. difference 0.6%

Jet Fuel production, Fischer-Tropsch:

Reaction:	15 CO	+	21 H ₂	--->	1 C ₁₁ H ₂₄	+	8 H ₂ O	+	0 H ₂ + 3 CO ₂
Energy [kJ]	4242	+	5038		7322		0		0 0

Heat out [kJ] 1958

AP Heat duty [Watt] 546

AP Energy [kJ] 1964

Rel. difference 0.3%

Ratio 1:1

Carbon input: 17

Steam methane reforming

Reaction:	8.5 CH ₄	+	8.5 CO ₂	+	0 H ₂ O	--->	17 CO	+	17 H ₂
Energy [kJ]	6801	+	0	+	0	--->	4808	+	4079

Heat out [kJ] -2085

AP Heat duty [Watt] -583

AP Energy [kJ] -2099

Rel. difference 0.7%

Jet Fuel production, Fischer-Tropsch:

Reaction:	17 CO	+	17 H ₂	--->	1 C ₁₁ H ₂₄	+	5 H ₂ O	+	0 H ₂	+	6 CO ₂
Energy [kJ]	4808	+	4079		6908		0		0		0

Heat out [kJ] 1979

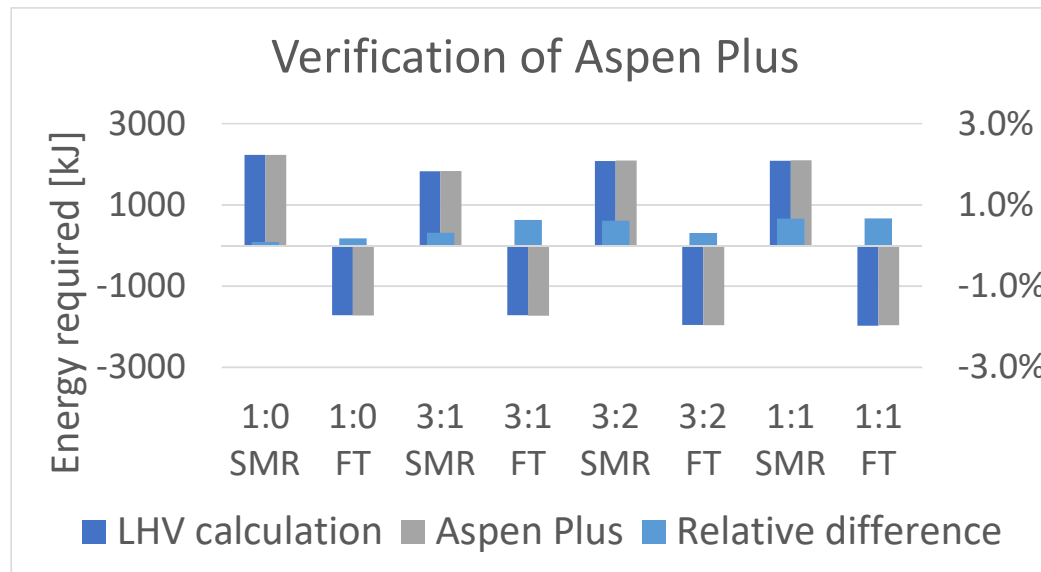
AP Heat duty [Watt] 546

AP Energy [kJ] 1966

Rel. difference 0.7%

Results

CH ₄ :CO ₂	1:0 SMR	1:0 FT	3:1 SMR	3:1 FT	3:2 SMR	3:2 FT	1:1 SMR	1:1 FT
LHV calculation	2227	-1721	1828	-1721	2079	-1958	2085	-1979
Aspen Plus	2228	-1724	1833	-1732	2092	-1964	2099	-1966
Relative difference	0.1%	0.2%	0.3%	0.6%	0.6%	0.3%	0.7%	0.7%



67

Prerequisites

Bionatural gas future potential	166.6666667
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BIOGAS

Results, Biogas

Inputs

	HOURLY	YEARLY	
	Amount [kmol/hour]	Amount [Mmol/year]	Recycled tail gas [PJ/year]
Methane consumption (CH4)			
SMR-heater	9058.9	79355.92	
SMR	13190.7	115550.18	
SUM	22249.6	194906.10	
ENERGY	[TJ/hour]	[PJ/year]	
SMR-heater		7.27	63.70
SUM heating		7.27	63.70
			41.82
SMR		10.59	92.75
SUM total methane		17.86	156.45
			134.57

Extra energy requirements

Name	SEPA
Heat duty [Watt]	0
Energy [TJ/hour]	0
Energy [PJ/year]	0

SUM SMR [PJ/year]	0.00
SUM FT [PJ/year]	0

Total energy input

	[TJ/hour]	[PJ/year]	
Total energy input		17.86	156.45
			Recycled tail gas [PJ/year]
			134.57

BIOGAS

Outputs

Steam methane reforming

BEFORE SEPARATION

	HOURLY [kmol/hour]	HOURLY [TJ/hour]		YEARLY [Mmol/year]	YEARLY [PJ/year]
HYDROGEN		34,532	8.35	302502.51	73.15
OXYGEN	trace				
NITROGEN					
WATER		22,593		197915.06	
CO		13,765	3.90	120581.72	34.12
CO2		7,103		62222.80	
METHANE		1,116	0.90	9779.13	7.85
UNDECANE	trace				
OCTANE	trace				
PROPANE	trace				
BUTANE	trace				
SUM		79,109.727	13.14	693001.21	115.13

AFTER SEPARATION

PRODUCTS	HOURLY [kmol/hour]	HOURLY [TJ/hour]		YEARLY [Mmol/year]	YEARLY [PJ/year]
HYDROGEN		28,992	7.01	253969.21	61.42
OXYGEN					
NITROGEN					
WATER				0.00	
CO		13,765	3.90	120581.72	34.12
CO2				0.00	
METHANE			0.00	0.00	0.00
UNDECANE					
OCTANE					
PROPANE					
BUTANE					

			BIOGAS		
SUM		42,756.955	10.91	374550.93	95.54
AFTER SEPARATION					
WASTE	HOURLY	HOURLY	YEARLY	YEARLY	
	[kmol/hour]	[TJ/hour]	[Mmol/year]	[PJ/year]	
HYDROGEN		5,540	1.34	48533.30	11.74
OXYGEN					
NITROGEN					
WATER		22,593		197915.06	
CO			0.00	0.00	0.00
CO2		7,103		62222.80	
METHANE		1,116	0.90	9779.13	7.85
UNDECANE					
OCTANE					
PROPANE					
BUTANE					
SUM		36,352.772	2.24	318450.28	19.59
Performance stats					
H2O input [kmol]		25000.00	Share:		
CH4 input [kmol]		13190.66	60%		
CO2 input [kmol]		8793.77	40%		
H2O:C		1.14			
H2 output [kmol]		34532.25			
CO output [kmol]		13765.04			
H2:CO		2.51			
CO2 output [kmol]		7103.06			
CO output [kmol]		13765.04			
CO2:CO		0.52			
CH4 output [kmol]		1116.34			

BIOGAS

Total molar output [kmol]	79109.73
CH4 mol%	1%
Carbon efficiency	63%

FISCHER-TROPSCH REACTOR

OUTPUT	[kmol/hour]	[Mmol/year]	C-atoms	Output molar %	Energy [TJ/hour]	Energy [PJ/year]	
HYDROGEN	1083.519				0.26	2.30	
OXYGEN							
NITROGEN							
WATER	12262.916						
CO							
CO2	338.109						
METHANE							
UNDECANE	720.49		6312	11	55%	4.98	43.60
OCTANE	584.45		5120	8	45%	2.97	25.98
PROPANE							
BUTANE							
Carbon efficiency		92%					
Total carbon efficiency		57%					

Energy output

	[TJ/hour]	[PJ/year]		
UNDECANE		4.98	43.60	
OCTANE		2.97	25.98	
Energy efficiency	SMR input	Product efficiency	Product efficiency with recycling	
UNDECANE		47%	28%	32%
OCTANE		28%	17%	19%
TOTAL		75%	44%	52%

BIOGAS

District heating output

District heating exchanger	DH1	DH2	DH3	SUM
Heat duty [W]	280235739	691285905	557621143	1529142787
Energy [TJ/hour]	1.00884866	2.488629258	2.007436115	5.504914033
Energy [PJ/year]	8.837514265	21.8003923	17.58514037	48.22304693
SUM SMR [PJ/year]	30.63790657			
SUM FT [PJ/year]	17.58514037			
	Total efficiency	DH Efficiency		
Energy efficiency incl. DH	75%	31%		
With methane recycling	88%	36%		

Heat exchangers

Name	DH1	DH2	DH3	ECO	EVA	FEEDSUP1	FEEDSUP2
Heat duty [Watt]	280235739	691285905	557621143	137787540	311471263	104907260	88670616.7
Energy [TJ/hour]	1.00884866	2.488629258	2.007436115	0.496035144	1.121296547	0.377666136	0.31921422
Energy [PJ/year]	8.837514265	21.8003923	17.58514037	4.345267861	9.82255775	3.308355351	2.796316568
					PRODHEAT	SG-HEAT	SUP
					41972172.1	38898832.5	161917223
SUM SMR [PJ/year]	27.93				0.15109982	0.140035797	0.582902003
SUM FT [PJ/year]	0				1.323634419	1.226713582	5.106221545

Losses

	[PJ/year]
Losses in reformer	10.69
Losses in Fischer-Tropsch reactor	6.08

BIOMETHANE

Prerequisites

Hydrogen

Lower heating value 241.8 kJ/mol ; MJ/kmol

CO

Lower heating value 283.0 kJ/mol ; MJ/kmol

Methane

Lower heating value 802.7 kJ/mol ; MJ/kmol

50.0 MJ/kg

Molar weight 16.0 kg/kmol;ton/Mmol

Undecane (C₁₁H₂₄)

11.0

Lower heating value 6908.0 kJ/mol ; MJ/kmol

44.2 MJ/kg

Molar weight 156.0 kg/kmol;ton/Mmol

Octane (C₈H₁₈)

8.0

Lower heating value 5074.9 kJ/mol ; MJ/kmol

44.4 MJ/kg

Molar weight 114.0 kg/kmol;ton/Mmol

Production and potential

Biogas 2017 [PJ] 11.158

Biogas potential [PJ] 50

Bionatural gas 2017 [PJ] 5.13

Bionatural gas potential [PJ] 80

Biogas future potential [PJ] 100

Bionatural gas future potential [P. 166.666667

Results, Biomethane

Inputs

	HOURLY	YEARLY	
	Amount [kmol/hour]	Amount [Mmol/year]	Recycled tail gas [PJ/year]
Methane consumption (CH4)			
SMR-heater	13116.5	114900.40	
SMR	21352.3	187046.16	
SUM	34468.8	301946.56	
ENERGY	[TJ/hour]	[PJ/year]	
SMR-heater		92.23	
SUM heating		92.23	3.10
SMR		150.14	
SUM total methane		242.37	153.24

Extra energy requirements

Name	SEPA
Heat duty [Watt]	0
Energy [TJ/hour]	0
Energy [PJ/year]	0
SUM SMR [PJ/year]	0.00
SUM FT [PJ/year]	0

Total energy input

	[TJ/hour]	[PJ/year]	Recycled tail gas [PJ/year]
Total energy input		27.67	153.24

BIOMETHANE

Outputs					
Steam methane reforming					
BEFORE SEPARATION					
	HOURLY	HOURLY		YEARLY	YEARLY
	[kmol/hour]	[TJ/hour]		[Mmol/year]	[PJ/year]
HYDROGEN		61,886	14.97	542123.40	131.10
OXYGEN	trace				
NITROGEN					
WATER		29,320		256844.50	
CO		13,765	3.90	120581.62	34.12
CO2		5,148		45094.64	
METHANE		2,439	1.96	21369.90	17.15
UNDECANE	trace				
OCTANE	trace				
PROPANE	trace				
BUTANE	trace				
SUM		112,558.682	20.82	986014.05	182.38
AFTER SEPARATION					
PRODUCTS	HOURLY	HOURLY		YEARLY	YEARLY
	[kmol/hour]	[TJ/hour]		[Mmol/year]	[PJ/year]
HYDROGEN		28,992	7.01	253969.00	61.42
OXYGEN					
NITROGEN					
WATER				0.00	
CO		13,765	3.90	120581.62	34.12
CO2				0.00	
METHANE			0.00	0.00	0.00
UNDECANE					
OCTANE					
PROPANE					
BUTANE					

BIOMETHANE					
SUM		42,756.920	10.91	374550.62	95.54
AFTER SEPARATION					
WASTE	HOURLY	HOURLY	YEARLY	YEARLY	
	[kmol/hour]	[TJ/hour]	[Mmol/year]	[PJ/year]	
HYDROGEN		32,894	7.95	288154.40	69.68
OXYGEN					
NITROGEN					
WATER		29,320		256844.50	
CO			0.00	0.00	0.00
CO2		5,148		45094.64	
METHANE		2,439	1.96	21369.90	17.15
UNDECANE					
OCTANE					
PROPANE					
BUTANE					
SUM		69,801.762	9.91	611463.44	86.84
Performance stats					
H2O input [kmol]		25000.00	Share:		
CH4 input [kmol]		21352.30	100%		
CO2 input [kmol]		0.00	0%		
H2O:C		1.17			
H2 output [kmol]		61886.23			
CO output [kmol]		13765.03			
H2:CO		4.50			
CO2 output [kmol]		5147.79			
CO output [kmol]		13765.03			
CO2:CO		0.37			
CH4 output [kmol]		2439.49			

BIOMETHANE

Total molar output [kmol]	112558.68
CH4 mol%	2%
Carbon efficiency	64%

FISCHER-TROPSCH REACTOR

OUTPUT	[kmol/hour]	[Mmol/year]	C-atoms	Output molar r	Energy [TJ/hou	Energy [PJ/year]	
HYDROGEN	1083.518				0.26	2.30	
OXYGEN							
NITROGEN							
WATER	12262.906						
CO							
CO2	338.108						
METHANE							
UNDECANE	720.49		6312	11	55%	4.98	43.60
OCTANE	584.45		5120	8	45%	2.97	25.98
PROPANE							
BUTANE							
Carbon efficiency		92%					
Total carbon efficiency		59%					

Energy output

	[TJ/hour]	[PJ/year]		
UNDECANE		4.98	43.60	
OCTANE		2.97	25.98	
Energy efficiency	SMR input	Product efficiency	Product efficiency with recycling	
UNDECANE		29%	18%	28%
OCTANE		17%	11%	17%
TOTAL		46%	29%	45%

BIOMETHANE

District heating output				
District heating exchanger	DH1	DH2	DH3	SUM
Heat duty [W]	361514893	947951148	557620686	1867086727
Energy [TJ/hour]	1.301453615	3.412624133	2.00743447	6.721512217
Energy [PJ/year]	11.40073367	29.8945874	17.58512595	58.88044702
SUM SMR [PJ/year]	41.29532107			
SUM FT [PJ/year]	17.58512595			
	Total efficiency	DH Efficiency		
Energy efficiency incl. DH	53%	24%		
With methane recycling	84%	38%		

Heat exchangers							
Name	DH1	DH2	DH3	ECO	EVA	FEEDSUP1	FEEDSUP2
Heat duty [Watt]	361514893	947951148	557620686	223042746	504192220	91736972.2	104699840
Energy [TJ/hour]	1.301453615	3.412624133	2.00743447	0.802953886	1.815091992	0.3302531	0.376919424
Energy [PJ/year]	11.40073367	29.8945874	17.58512595	7.033876038	15.90020585	2.893017155	3.301814154
					PRODHEAT	SG-HEAT	SUP
					41972138.5	34607774	262102524
SUM SMR [PJ/year]	39.81				0.151099699	0.124587986	0.943569086
SUM FT [PJ/year]	0				1.32363336	1.091390761	8.265665197

Losses	
	[PJ/year]
Losses in reformer	18.70
Losses in Fischer-Tropsch reactor	6.08

Prerequisites

Hydrogen

Lower heating value 241.8 kJ/mol ; MJ/kmol

CO

Lower heating value 283.0 kJ/mol ; MJ/kmol

Methane

Lower heating value 802.7 kJ/mol ; MJ/kmol

50.0 MJ/kg

Molar weight 16.0 kg/kmol;ton/Mmol

Undecane (C₁₁H₂₄)

11.0

Lower heating value 6908.0 kJ/mol ; MJ/kmol

44.2 MJ/kg

Molar weight 156.0 kg/kmol;ton/Mmol

Octane (C₈H₁₈)

8.0

Lower heating value 5074.9 kJ/mol ; MJ/kmol

44.4 MJ/kg

Molar weight 114.0 kg/kmol;ton/Mmol

Production and potential

Biogas 2017 [PJ] 11.158

Biogas potential [PJ] 50

Bionatural gas 2017 [PJ] 5.13

Bionatural gas potential [PJ] 80

Biogas future potential [PJ] 100

Bionatural gas future potential [P. 166.666667

Results, Biomethane and CO2

Inputs

	HOURLY	YEARLY	
	Amount [kmol/hour]	Amount [Mmol/year]	Recycled tail gas [PJ/year]
Methane consumption (CH4)			
SMR-heater	8250.8	72277.14	
SMR	11523.3	100944.31	
SUM	19774.1	173221.45	
ENERGY	[TJ/hour]	[PJ/year]	
SMR-heater		6.62	58.02
SUM heating		6.62	58.02
			49.86
SMR		9.25	81.03
SUM total methane		15.87	139.04
			130.88

Extra energy requirements

Name	SEPA
Heat duty [Watt]	0
Energy [TJ/hour]	0
Energy [PJ/year]	0

SUM SMR [PJ/year]	0.00
SUM FT [PJ/year]	0

Total energy input

	[TJ/hour]	[PJ/year]	
Total energy input		15.87	139.04
			Recycled tail gas [PJ/year]
			130.88

BIOMETHANE AND CO2

Outputs

Steam methane reforming

BEFORE SEPARATION

	HOURLY [kmol/hour]	HOURLY [TJ/hour]		YEARLY [Mmol/year]	YEARLY [PJ/year]
HYDROGEN		28,992	7.01	253969.43	61.42
OXYGEN	trace				
NITROGEN					
WATER		21,195		185666.84	
CO		13,765	3.90	120581.56	34.12
CO2		7,935		69511.13	
METHANE		834	0.67	7306.57	5.86
UNDECANE	trace				
OCTANE	trace				
PROPANE	trace				
BUTANE	trace				
SUM		72,720.950	11.58	637035.52	101.40

AFTER SEPARATION

PRODUCTS	HOURLY [kmol/hour]	HOURLY [TJ/hour]		YEARLY [Mmol/year]	YEARLY [PJ/year]
HYDROGEN		28,992	7.01	253969.43	61.42
OXYGEN					
NITROGEN					
WATER				0.00	
CO		13,765	3.90	120581.56	34.12
CO2				0.00	
METHANE			0.00	0.00	0.00
UNDECANE					
OCTANE					
PROPANE					
BUTANE					

BIOMETHANE AND CO2

SUM	42,756.962	10.91	374550.99	95.54
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AFTER SEPARATION

WASTE	HOURLY [kmol/hour]	HOURLY [TJ/hour]	YEARLY [Mmol/year]	YEARLY [PJ/year]
HYDROGEN		0	0.00	0.00
OXYGEN				
NITROGEN				
WATER	21,195		185666.84	
CO			0.00	0.00
CO2	7,935		69511.13	
METHANE	834		0.67	7306.57
UNDECANE				
OCTANE				
PROPANE				
BUTANE				
SUM	29,963.988		0.67	262484.53

Performance stats

H2O input [kmol]	25000.00	Share:
CH4 input [kmol]	11523.32	
CO2 input [kmol]	11010.84	
H2O:C	1.11	
H2 output [kmol]	28991.94	
CO output [kmol]	13765.02	
H2:CO	2.11	
CO2 output [kmol]	7935.06	
CO output [kmol]	13765.02	
CO2:CO	0.58	
CH4 output [kmol]	834.08	

BIOMETHANE AND CO2

Total molar output [kmol]	72720.95
CH4 mol%	1%
Carbon efficiency	61%

FISCHER-TROPSCH REACTOR

OUTPUT	[kmol/hour]	[Mmol/year]	C-atoms	Output molar %	Energy [TJ/hour]	Energy [PJ/year]	
HYDROGEN	1083.535				0.26	2.30	
OXYGEN							
NITROGEN							
WATER	12262.927						
CO							
CO2	338.095						
METHANE							
UNDECANE	720.49		6312	11	55%	4.98	43.60
OCTANE	584.45		5120	8	45%	2.97	25.98
PROPANE							
BUTANE							
Carbon efficiency		92%					
Total carbon efficiency		56%					

Energy output

	[TJ/hour]	[PJ/year]		
UNDECANE		4.98	43.60	
OCTANE		2.97	25.98	
Energy efficiency	SMR input	Product efficiency	Product efficiency with recycling	
UNDECANE		54%	31%	33%
OCTANE		32%	19%	20%
TOTAL		86%	50%	53%

BIOMETHANE AND CO2

District heating output

District heating exchanger	DH1	DH2	DH3	SUM
Heat duty [W]	263087974	639499602	557620873	1460208449
Energy [TJ/hour]	0.947116706	2.302198567	2.007435143	5.256750416
Energy [PJ/year]	8.296742348	20.16725945	17.58513185	46.04913365
SUM SMR [PJ/year]	28.4640018			
SUM FT [PJ/year]	17.58513185			
	Total efficiency	DH Efficiency		
Energy efficiency incl. DH	83%	33%		
With methane recycling	88%	35%		

Heat exchangers

Name	DH1	DH2	DH3	ECO	EVA	FEEDSUP1	FEEDSUP2
Heat duty [Watt]	263087974	639499602	557620873	120370805	272100414	110437624	86003122.4
Energy [TJ/hour]	0.947116706	2.302198567	2.007435143	0.433334898	0.97956149	0.397575446	0.309611241
Energy [PJ/year]	8.296742348	20.16725945	17.58513185	3.796013706	8.580958656	3.48276091	2.712194468
					PRODHEAT	SG-HEAT	SUP
					41972177.2	39770162.9	141450428
SUM SMR [PJ/year]	25.61				0.151099838	0.143172586	0.509221541
SUM FT [PJ/year]	0				1.32363458	1.254191857	4.460780697

Losses

	[PJ/year]
Losses in reformer	9.18
Losses in Fischer-Tropsch reactor	6.08