

Methanol Via Co-gasification of Black Liquor and By-products from Biofuel Production

A Techno-economic Analysis

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Abstract

Previous studies and pilot scale plants have shown that gasification of black liquor for biofuel production has the potential to significantly contribute to a fossil independent vehicle fleet Sweden's 2030 targets. To improve the plants profitability, increase the production, and competitiveness towards fossil fuels, co-gasification is of interest since black liquor availability is limited to the production of paper pulp.

The main objective of this project was to perform a techno-economic analysis on how co-gasification of black liquor and either crude glycerol or fermentation residues to methanol would affect the production rate, profitability, and efficiency of a full scale integrated biofuel production plant.

The results show that from a technical and economic perspective the best choice is co-gasification of black liquor and 50 %wt. of crude glycerol. This will improve the conversion efficiency from feedstock to methanol from 51 % (using black liquor only) to 62 %, improve the biofuel production rate more than 240 %, and give an internal rate of return of 21.9 % compared to 11.5 % for the reference case of using black liquor. Co-gasification of black liquor and 50 %wt. of crude glycerol also have a good chance to make a price match with fossil derived methanol and still maintain a reasonable internal rate of return of 11.3 % for the investment.

The highest reasonable cost for crude glycerol without receiving a lower internal rate of return than for the reference case of black liquor gasification is 50.5 €/MWh. This can be compared to the cost used in this thesis of 20 €/MWh.

With an investment cost of 583 M€, an integrated biofuel plant with methanol production via co-gasification of black liquor and 50 %wt. of crude glycerol is considered a good investment. This also gives the pulp mill another commodity on the market, which probably will lower the risk of having a poor financial statement.

The price for methanol is the key factor to the investment's profitability. This demands thorough investigations and market studies to ensure that the selling price for methanol is competitive

Sammanfattning

Tidigare studier och pilotskalanläggningar har visat att förgasning av svartlut till biodrivmedel är en tekniskt genomförbar metod till att bidra till en fossilfri fordonsflotta enligt Sveriges 2030-mål. För att öka lönsamheten, produktionen av biodrivmedel samt dess konkurrenskraft mot fossila drivmedel krävs samförgasning, då tillgången till svartlut är begränsad till produktionen av pappersmassa.

Målet med detta projekt var att genomföra en tekno-ekonomisk analys om hur samförgasning av svartlut tillsammans med antingen råglycerol eller fermenteringsrester till metanol hade för möjlighet att påverka en fullskalig integrerad biodrivmedelsanläggnings produktion, lönsamhet och effektivitet.

Sett ur tekniskt och ekonomiskt perspektiv är den bästa lösningen en samförgasning av svartlut tillsammans med en 50 % av totala massan inblandning av råglycerol. Denna blandning förbättrar konverteringseffektiviteten från råmaterial till metanol från 51 % med endast svartlutsförgasning till 62 %, ger en produktionsökning av metanol med drygt 240 %, samt att en internränta för investeringen på 21,9 % erhålls jämfört med 11,5 % i referensfallet med endast svartlut som råmaterial. En samförgasning av svartlut och 50 % råglycerol har dessutom goda möjligheter att prismatcha fossilt utvunnen metanol och samtidigt erhålla en god återbetalning på investeringen, då med en internränta på 11,3 %.

Maximala priset som anses vara möjligt att betala för råglycerolen utan att erhålla en sämre internränta än för referensfallet med svartlutsförgasning är 50,5 €/MWh, detta kan jämföras med pris som använts i detta examensarbete på 20 €/MWh.

Med en investeringskostnad på 583 miljoner euro anses det att en integrerad biodrivmedelsanläggning med produktion av metanol via samförgasning av svartlut och 50 % råglycerol är en bra investering. Detta ger pappersmassabruket ytterligare en produkt på marknaden, vilket mest sannolikt leder till att minska risken för ett dåligt bokslut.

Priset på metanol är den störst avgörande faktor huruvida en investering är lönsam eller inte, vilket kräver noggrant utförda marknadsstudier för att säkerställa att rätt pris fås vid försäljning av metanolen.

Glossary and variable list

Table 1. Abbreviations and their actual meaning that will be used in the report.

Abbreviation	Description
AGR	Acid gas removal
AGE	Acid gas enrichment
ASU	Air separation unit
BAT	Best available technology
BL	Black liquor
BLG	Black liquor gasification
BFP	Biofuel plant
BOP	Balance of plant
CAPEX	Capital expenses
CDE	Carbon dioxide equivalents
CECPI	Chemical engineering price index
CG	Crude glycerol
CHP	Combined heat and power
CPI	Consumer price index
DS	Dry substance
EFG	Entrained flow gasifier
FAME	Fatty acid methyl ester
FFA	Free fatty acids
FR	Fermentation residue
FW	Feedwater
GC	Gas cooling
GDP	Gross domestic product
GHG	Greenhouse gas
HEX	Heat exchanger
HHV	Higher heating value
HW	Hot water
IPCC	Intergovernmental Panel on Climate Change
IRR	Internal rate of return
KPP	Kraft pulp process
LPS	Low pressure steam
MONG	Matter organic non-glycerol
MPS	Medium pressure steam
MSY	Methanol synthesis
MUW	Make-up water
NPV	Net present value
OPEX	Operating expenses
PaPM	Pulp and paper mill
PBT	Pay-back time
PDU	Process distillation unit
PF	Process fuel
PM	Pulp mill
PS	Process steam
SF-LHV	Sulphur-free lower heating value
SG	Synthesis gas
SNG	Synthetic natural gas
WLA	White liquor adsorption

Table 2. Variables used in equations in the thesis with unit and description.

Variable	Unit	Description
n	mole	Amount of moles
H	MJ/kg	Heating value
ΔH_f^0	kJ/mole	Heat of formation
m	kg/kg	Mass fraction
M	%	Moisture content
C	€	Cost
S	MW	Equipment capacity
k	-	Scaling coefficient
d	%	Discount rate
a	-	Annuity factor
t	Years	Technical life expectancy
D	€/year	Annual depreciation
c	€	Cash flow
R	€	Revenue
r	%	Real interest rate
i	%	Internal rate of return

Table 3. Suffixes for variables in Table 2 with description.

Suffix	Description
f	Formation
i	Component i
j	Component j
H	Hydrogen
S	Sulphur
0	Initial state
y	Specific year

Table of contents

1	Introduction.....	10
1.1	International climate and energy goals.....	11
1.2	Swedish CO ₂ -emission and energy goals.....	12
1.3	Renewable transport fuels	12
1.4	Biofuel production via black liquor gasification	14
1.5	Ethanol pricing.....	16
2	Aim and objectives	17
2.1	Scope	17
2.2	Case descriptions.....	18
2.3	Fuel feedstock	18
2.3.1	Black liquor	19
2.3.2	Crude glycerol.....	19
2.3.3	Fermentation residues	21
3	Theory.....	21
3.1	Crude glycerol.....	21
3.2	Heating value.....	22
3.3	Equipment and labour costs.....	22
3.4	Straight payback time.....	23
3.5	Net present value	23
3.6	Internal rate of return	24
4	Method	24
4.1	Aspen® simulation model.....	24
4.1.1	Synthesis gas composition.....	25
	Black liquor composition	26
	Crude glycerol composition.....	26
	Fermentation residue composition	27
	Heating values	27
4.2	Techno-economic analysis	28
4.2.1	Material and energy flow costs and efficiency.....	29
4.2.2	Equipment and labour costs.....	31
4.2.3	Summary.....	33
4.2.4	Sensitivity analysis.....	33
5	Results and analysis.....	33
5.1	Aspen® simulation model.....	33

5.1.1	Synthesis gas composition.....	38
5.2	Techno-economic analysis	41
5.2.1	Material and energy flow costs and efficiency.....	41
5.2.2	Equipment and labour costs.....	41
5.2.3	Summary.....	42
5.2.4	Sensitivity analysis.....	47
6	Conclusion	51
7	Future work	52
7.1	Issues	52
7.2	Improvements	53
8	References.....	54
9	Appendix.....	I
9.1	Appendix 1 – Aspen® modelling settings	I
9.2	Appendix 2 – Feedstock composition calculation sheet	IV
9.3	Appendix 3 – Boiler calculations to determine biomass flow	V
9.4	Appendix 4 – Project time plan	VI
9.5	Appendix 5 – Exemplary calculation sheet for economic analysis.....	VII

1 Introduction

The Intergovernmental Panel on Climate Change (**IPCC**) is the collected international knowledge and science regarding climate change. The IPCC provides scientific information of climate change to governments and legislators, such as impacts, future risks, and options for adaptation. This information can be used to help establish for instance climate-related policies and regulations of greenhouse gas (**GHG**) emissions. There are no mandatory implementations or measures provided by the IPCC, only predictions of climate change, the risks they pose, and recommendations to respond to a possible impending climate catastrophe.

The last 30 years (1983 to 2012) is determined to be very likely the warmest period in 800 years in the Northern Hemisphere. Globally the average land and sea temperature has shown an increase of 0.85°C between the periods 1880 to 2012. More than half of the observed increase in global surface temperature from post-world war II (1950th) to this day is most certainly caused by anthropogenic release of GHG emission into the atmosphere (IPCC, 2014).

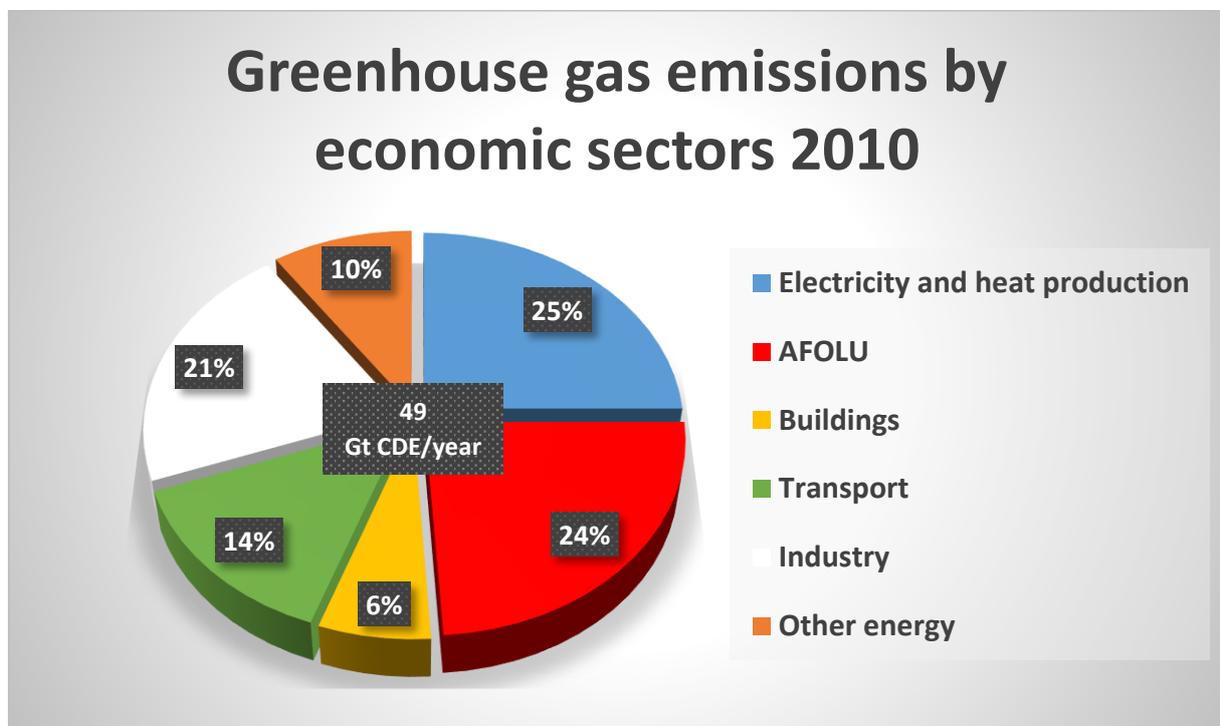


Figure 1. Annual anthropogenic originated GHG emission in different economic sectors and their contribution in 2010. AFOLU is short for agriculture, forestry, and other land use, which includes forest fires, peat fires, and peat decay (IPCC, 2014).

As seen in Figure 1, from the total GHG emission of 49 Gigatonne of carbon dioxide equivalents (**CDE**) that originates from anthropogenic emissions in the economic sector, 14 % of these derive from internal combustion engines in the transport sector. A majority (~97 %) originates from fossil sources according to figures from 2010 (IPCC, 2014).

To quote the IPCC in the Fifth assessment report:

Anthropogenic greenhouse gas emissions have increased since the pre-industrial era, driven largely by economic and population growth, and are now higher than ever. This has led to atmospheric concentrations of carbon dioxide, methane and nitrous oxide that are unprecedented in at least the last 800 000 years. Their effects, together with those of other anthropogenic drivers, have been detected throughout the climate system and are extremely likely to have been the dominant cause of the observed warming since the mid-20th century (IPCC, 2014).

In order to reduce the amount of GHG emitted into the atmosphere, measures has to be taken to avoid a severe and irreversible impact on the environment, eco-systems, people as well as economic stability and growth (IPCC, 2014). To address the GHG emissions from transport, one measure would be to switch the current fossil sources to renewable ones. The main advantage with fuel deriving from biological sources is that its carbon contents comes from photosynthesis and growing of the feedstock. This carbon ends up in the fuel and is simply re-introduced into the atmosphere, where fossil carbon that originates from outside the atmosphere system boundary increase the total concentration of carbon. This will be further described in section 1.3.

1.1 International climate and energy goals

A strategy for a smart, sustainable and inclusive growth in the European Union (EU) was made in 2010, also known as “Europe 2020 Strategy”. Under this strategy, three goals are set to address the climate issue.

The first goal is to reduce the amount of greenhouse gas (GHG) emissions by 20 % when compared to 1990 levels. This will achieved through incentives, taxes, new technology, increased efficiency, and replacement of fossil based fuels (Directive 2009/28/EC, 2009).

In order to reduce the dependence on foreign fossil fuel and derivatives as well as GHG emissions, a renewable energy target has been set in the EU to increase the market share for renewable energy sources such as biofuels, wind and solar power. The goals are set to achieve an overall energy production of 20 % from renewable sources, and a minimum target share of 10 % for biofuels compared to diesel and petrol consumption before year 2020. According to (European commission, 2010) and the EU 2020, an increased share of renewable energy production will reduce import of foreign oil and gas and is estimated to generate savings of € 60 billion, create more than 600 000 new jobs and add 0.6-0.8 % to GDP by 2020.

The final climate goal in the EU 2020 strategy is a 20 % improvement in the EU's energy efficiency, which will be achieved through building regulation, promotions, taxation, and subsidies (European commission, 2010).

To meet the effects of a climate change, an environment quality goal was set with a temperature and concentration goal. The global mean temperature should not increase greater than 2°C above the preindustrial temperature. The concentration of 400 ppm of CDE should also not be exceeded and stabilized at this level (Ministry of the Environment Sweden, 2014).

1.2 Swedish CO₂-emission and energy goals

The national goals set by the Swedish government aims to reduce the amount of GHG emissions by 40 % before the year 2020 when compared to the levels from the year 1990. This will end up in a reduced emission rate of 20 million tonnes of CDE on annual basis. The climate and energy goals are set for the sector which is not subjected to the emissions trading system. A vision is also set for Sweden; before the year 2030 the entire vehicle fleet should be independent from fossil fuels and by 2050 Sweden has no net emissions of GHG into the atmosphere (Prop., 2009/10:155).

In Sweden, the energy used in the transport sector is according to figures from 2012 around 90.2 TWh. Of this the renewable share of transport fuels ends up in 15.6 % if calculated in accordance with the Renewable Energy Directive. Looking at the amount of renewables in the transport sector based on energy content, the share ends up in 9.7 % (Swedish Energy Agency, 2014).

1.3 Renewable transport fuels

A variety of renewable transport fuels are available; some are currently on the market in substantial production scale, some that is under development, and others that has not yet acquired a large enough market share to become an economic viable alternative.

Ethanol is mainly produced through fermentation of starch, sugar or cellulose-rich materials (Svebio, n.d.). It is a liquid fuel that used to either replace or act as an additive to fossil petrol fuel in spark ignition engines.

By utilizing bacterial biological breakdown of organic materials, also known as anaerobic digestion, biogas can be produced. In a deoxygenated environment a synthetic natural gas (**SNG**) is produced, cleaned and can be used in vehicles either made for this type of fuel, or converted petrol engines (Svebio, n.d.). Diesel engines can also be converted to use this type of fuel, but requires either changing the ignition system or use diesel as an ignition fuel for the biogas (methane diesel) (SMP Svensk Maskinprovning AB, 2011).

Feedstock for biodiesel production generally consists of vegetable oils and uses a technique called transesterification, which uses alcohol and catalysts to separate the organic group of an ester. Generally biodiesel can be used in a regular diesel engine without or with minor modification (Biodiesel - America's Advanced Biofuel, 2012).

The so-called “second generation” biofuels is mainly characterised by their CO₂-reduction capabilities (see Figure 3) and is mainly produced from lignocellulosic biomass that is converted into various fuels, such as ethanol, dimethyl ether (**DME**), methanol, Fischer-Tropsch diesel, and hydrogen (Svebio, n.d.). Compared to “first generation” biofuels, feedstock does not compete with crops designated for food over arable land.

There are a variety of methods to produce second generation biofuels, such as thermochemical treatment or gasification into a synthesis gas (**SG**), fermentation, or hydrogenation of oils. Some commonly used fuel feedstocks are household/industrial waste, forestry residues, straw, and energy crops (switchgrass). Depending on the desired biofuel, different production and feedstocks are used, which is shown in Table 4. Observe that this is their commonly used treatment, however any of the feedstock material might be used in any of the methods above, with different results and efficiency. Lignocellulosic material needs to undergo pre-treatment prior to fermentation in order to ferment cellulose to ethanol or butanol.

Table 4. Second generation feedstocks and their commonly utilized method for producing biofuels (ÉcoRessources Consultants, 2012).

	Thermochemical treatment/ gasification	Biochemical treatment/ fermentation	Hydrogenation
Household/industrial waste	Yes	Yes	Yes
Forestry residues	Yes	Yes	
Straw	Yes	Yes	
Energy crops	Yes	Yes	

In order to determine how well different fuels and feedstocks contribute to an environmentally sustainable transport fuel, rules have been established by the EU to calculate the GHG net emission reduction. The savings are based on the amount of GHG emission the suggested fuel would have in comparison to its fossil derivative, and is presented in Figure 2 (Directive 2009/28/EC, 2009).

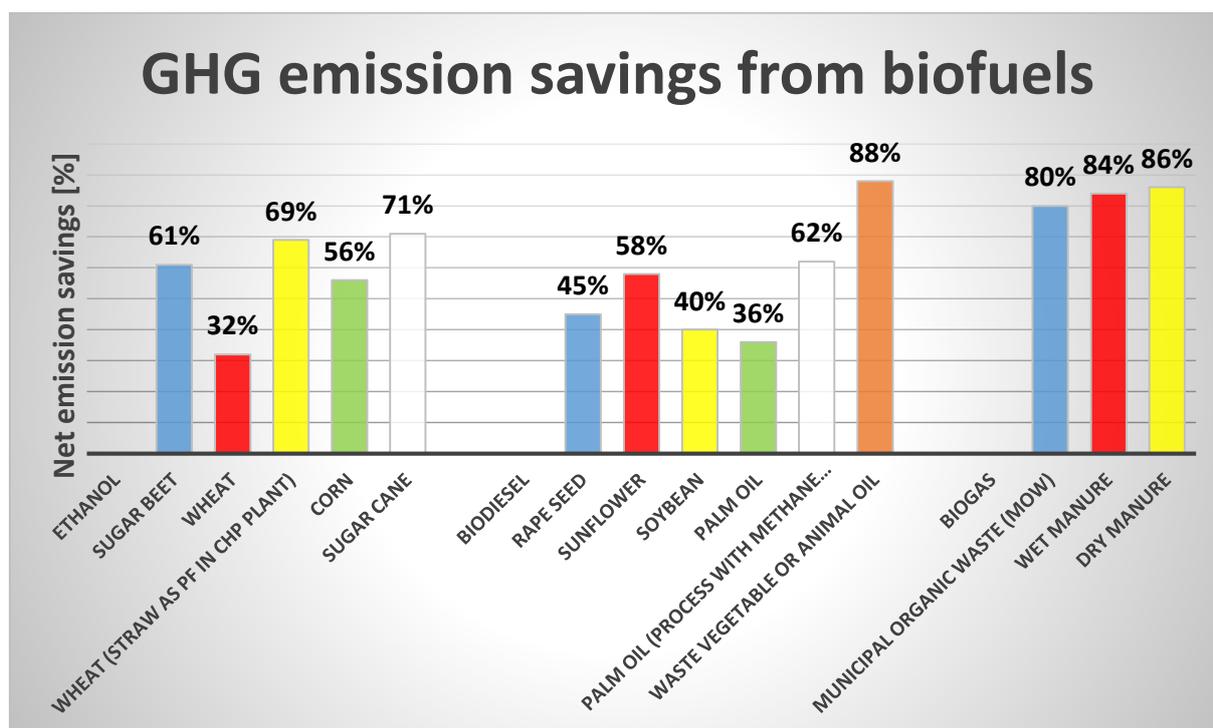


Figure 2. Typical values of net emission savings for various biofuels depending on fuel type and feedstock, assuming no carbon net emission from land-usage (Directive 2009/28/EC, 2009).

Some renewable fuels that are not or the market or has a negligible contribution, has also been evaluated. Their potential for GHG emission savings is shown in Figure 3.

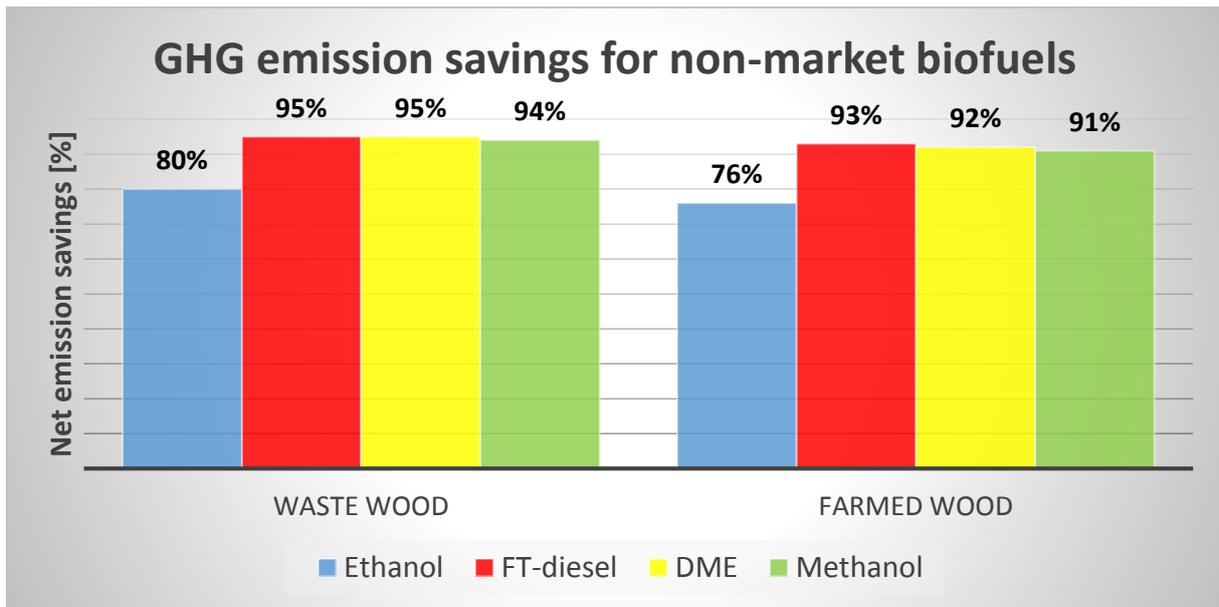


Figure 3. Estimated typical GHG emission savings depending on fuel type and feedstock, with no net carbon emission from land changing for feedstock (Directive 2009/28/EC, 2009).

1.4 Biofuel production via black liquor gasification

As previously described the general procedure to use BL is for combustion, which in turn will recover the cooking chemicals as well as produce heat and steam which is necessary for various processes for a paper mill (PM).

Another option to recover the cooking chemicals is to replace the recovery boiler with a gasification unit, which operates in sub-stoichiometric conditions. In an entrained flow gasifier (EFG), the black liquor is pre-heated for a lower viscosity and a finer dispersion of particles and then injected into the gasifier through a top mounted nozzle together with a determined oxygen-to-fuel ratio of pure oxygen (99.5 %vol.). The EFG temperature and pressure varies, but the EFG developed by Chemrec is operating around 1000°C and at pressures of 30 bar. In these conditions, the organic compounds become volatile gases, such as carbon monoxide, methane, and hydrogen, while the inorganic compounds becomes an ash melt. Both the gas and the melt is cooled in a quench process with water, where the melt solidifies and drops to the bottom of the EFG. This is dissolved in a weak wash and exits as “green liquor”. After causticizing in a lime kiln the recovered cooking chemicals are returned to the cooking process as “white liquor”. For graphical illustration of the process, see Figure 4. The output gas mixture goes through a secondary quench where it is generally referred to as “moist uncooled syngas” (Ekblom, Lindblom, Berglin, & Ahlvik, 2003).

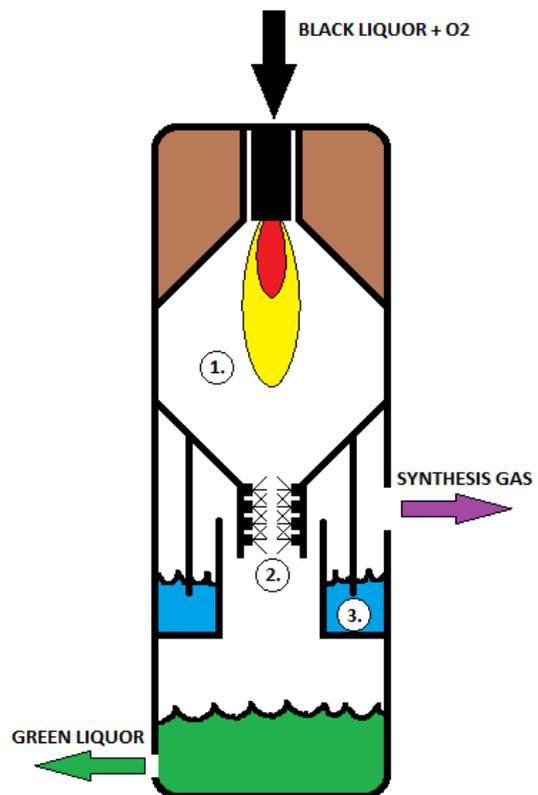


Figure 4. Cross section illustration of BLG in an EFG. Numbered sections: 1) Gasification zone; 2) Primary quench zone; 3) Secondary quench zone.

The output SG contains a high concentration of water (~73 % of volume) and a temperature of about 220°C. The gas is passed through a heat exchanger (**HEX**) where it is cooled to 150°C. The cold side of the HEX generates low pressure steam (**LPS**) that is used in mill/biofuel plant (**BFP**) processes. The water condensed from the SG is reused in the quench section of the EFG. A final HEX cools the SG to 40°C, which gives a dry SG with a water content less than 0.3 % of volume. The cooling agent in the final HEX is water, mainly for production of hot water (**HW**) (Ekbom et al., 2003).

To obtain a desired ratio between hydrogen and carbon monoxide, half of the SG is run through a water-gas shift (**WGS**) process, where process steam (**PS**) is used. In the WGS process, the ratio between $\frac{H_2-CO_2}{CO+CO_2}$ is shifted so that it is close to 2 (Ekbom et al., 2003). The water-gas shift process is an equilibrium reaction which is illustrated in equation (1-1).



In order to remove the acid gas content in the SG in form of carbon dioxide and hydrogen sulphide, a process using an acid gas removal (**AGR**) unit is used. There are a variety of AGR units on the market which separates these components with different methods, such as amine solvents, pressure swing adsorption, cold methanol solvents (Rectisol™), or a mixture of DME and polyethylene glycol (Selexol™). Results of outgoing SG composition depending on method and unit will vary (Korens, Simbeck, & Wilhelm, 2002). The main advantage of choosing a Rectisol™ or Selexol™ as an AGR is the ability to remove both H₂S and CO₂ to very low concentrations. The difference between these two units other than the solvent used in the process is that Selexol™ operates at a lower pressure, usually around 2-14 bar and has a greater selectivity ratio than a Rectisol™, which operates around 25-60 bar. However, the energy consumption of the Rectisol™ is much less than that of the Selexol™ unit and has a deeper sulphur removal (Korens et al., 2002) (Atsonios, Christodoulou, Koytsoumpa, Panopoulos, & Kakaras, 2013).

The shifted and cleaned SG goes through synthesis and depending on the desired transport fuel, method will vary. For methanol, the so-called “sweet” SG is compressed to pressures ranging from 80 to 110 bar (a) before undergoing synthesis (Atsonios et al., 2013). In this step there is a significant amount of heat that needs to be removed, where in this case PS is produced for the WGS process and excess steam is brought down to medium pressure steam (**MPS**) levels. The synthesis loop produces methanol based on the simplified equilibrium reactions of equation (1-1), (1-2) and (1-3).



The gas is cooled and condensed, where gas still unconverted into methanol is recycled into the synthesis loop. A portion of the recycled gas is taken out as “tail gas”, which could be used as a supplementary fuel for pulp mill (**PM**) processes such as the lime kiln. A final step to remove any carbon dioxide and excess water contents to achieve a grade AA (~99.85 %) methanol will be necessary (Ekbom et al., 2003). A block flow diagram of a BFP with pre-treatment and methanol synthesis (**MSY**) is presented in Figure 5.

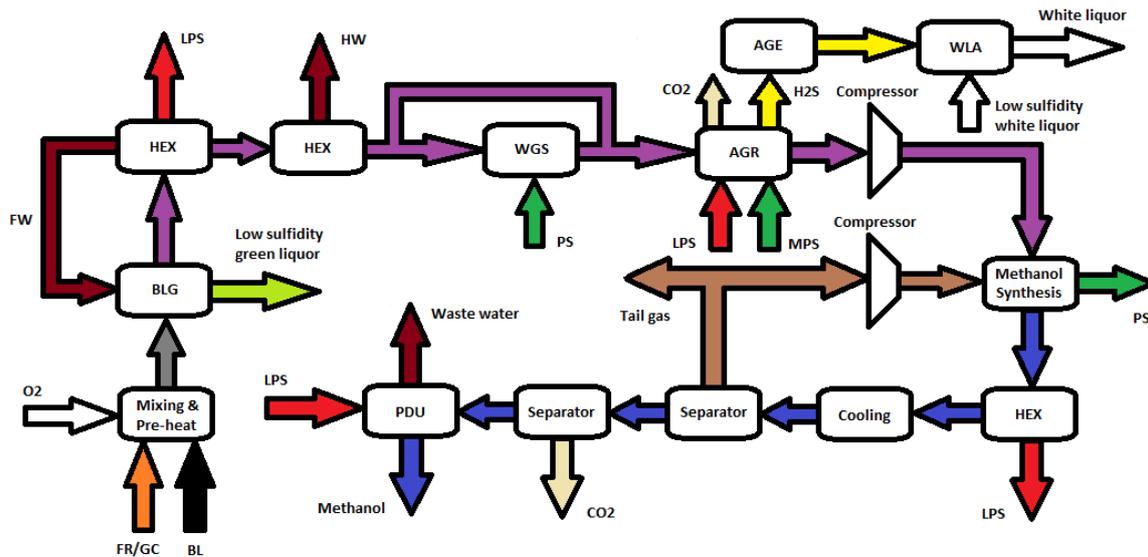


Figure 5. Block flow diagram of project BFP with material flows, excluding work and water into HEX for steam/HW production. Purple path illustrate flow of SG, while blue illustrate flow of methanol.

In order to make up for the steam required for the PM and BFP processes that used to be produced in the recovery boiler, a separate boiler for combustion of biomass will be necessary. A substantial amount of steam is produced in the BFP process, which in term generates a lower steam production in the boiler, lower electric power production, and thereby an increased import of electric power. A separate biomass boiler will also generate higher import of biomass on site for combustion. How the BFP is integrated with the PM can be seen in Figure 6.

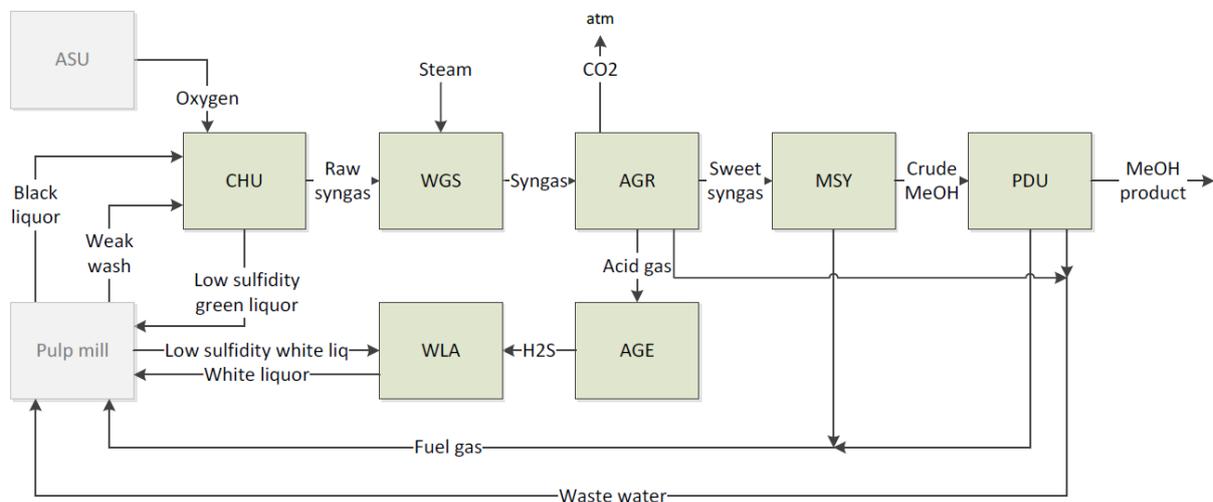


Figure 6. A simplified flowsheet illustration of a BFP for methanol production integrated with a PM (Andersson, Lundgren, & Furusjö, 2013).

1.5 Ethanol pricing

Like most types of commodities, prices fluctuates depending on market situations such as competitive price dumping or change in supply and demand. For ethanol sold in Europe there are two types of pricing, T1 and T2, where T1 ethanol is imported and T2 is ethanol produced in Europe or in countries that has a duty-free policy towards EU. In Figure 7 the fluctuation of T2 over a three year period is shown.

Rotterdam, fuel ethanol, FOB T2 [EUR/m³]



Figure 7. Price development of T2 ethanol from the port in Rotterdam between May 2012 and May 2015 in €/m³ (F.O. Lichts, 2015). FOB is a term that stands for Free on Board, which means that no transportation or insurance costs are included in the ethanol price.

2 Aim and objectives

By utilizing various by-products and blend them with BL to increase methanol production, a reliable and substantial source for biofuel production might be possible.

A pulp and paper mill (**PaPM**) with biofuel as an additional product at the market has the opportunity to increase revenues and lower the risk of having a poor financial statement with shifting market prices on pulp and paper products.

Production of biodiesel (RME) results in a by-product in form of crude glycerol (**CG**) fermentation of agricultural or forest residues to produce ethanol leaves a lignin rich unconverted fermentation residue (**FR**). By utilizing any of these two by-products in combination with black liquor (**BL**), an increase of methanol production might be achieved when availability of BL is limited. The aim of this work is to assess the techno-economic feasibility of the new technology to improve biofuel production efficiency. The main objectives of this study are to

- Make a draft design of commercial scale BLG process and methanol production, with additional feedstock supply chain
- Model and obtain mass and energy balance for the commercial scale design
- Estimate investment cost and profitability of an integrated BFP for different cases, see Case descriptions.

2.1 Scope

The project will be a strict techno-economic study which will focus on the production of methanol. Closing the material and energy balance and evaluating the added values will be the main subjects of the project. No technical experiments will be performed. All investment cost estimations will focus

on using best available technology (**BAT**). The techno-economic analysis will be a case study performed on data from Vallvik PM.

2.2 Case descriptions

The modelling will be based on different cases that will be techno-economically evaluated, see Table 5.

Table 5. Settings for the different cases in the project, where the "X" marks the case number conditions, "Low, medium, high" defines the amount of additional feedstock of the total weight in the BLG blend. Detailed description of blend quantity can be seen in Table 6.

Case no.	BLG	FR	CG	Low	Medium	High
1	X					
2	X	X		X		
3	X	X			X	
4	X	X				X
5	X		X	X		
6	X		X		X	
7	X		X			X

Due to varying mixing properties of additional feedstocks, the ratio between these and BL will vary for the different blend cases. Problems that may occur when mixing BL and additional feedstock is the formation of larger particles which might pose a problem in a gasification or combustion process. In consensus, mass ratios which are assumed to not cause any problem due to feedstock mixing can be seen in Table 6 (Häggström, 2015). While previous studies using pyrolysis oil up to 30 %wt. as additional feedstock showed no apparent problem in the catalytic effect (Andersson, Lundgren, & Furusjö, 2013), studies also shows for mixtures up to 40 %wt. of glycerol no significant change in char reactivity (Häggström, 2015). This thesis has not taken into consideration any reduction in the catalytic effect of the BL and additional fuel feedstock mixing when simulations was performed.

Table 6. Weight percentage distribution of the chosen feedstock at different blend cases, units in dry substance (DS).

Feedstock	Low [%wt.]	Medium [%wt.]	High [%wt.]
Fermentation residue	10	20	30
Crude glycerol	10	30	50

2.3 Fuel feedstock

In this section, some of the materials of interest for biofuel production will be described shortly. All of the fuel feedstock below is what would be considered biological waste or by-products that could be utilized for biofuel production, thus solving the issue of disposal as well as being a renewable feedstock.

According to Svebio (2008), utilizing the entire supply of BL in PaPM has the potential to replace 26 TWh of transport fuel, which is comparable with nearly two thirds of the entire diesel consumption in Sweden. According to Naturskyddsforeningen (2014) the potential for using waste wood material consisting of branches and stubs (GROT) gathered after foraging can increase to figures between 24.1 and 53.4 TWh, depending on which case is used. This can be compared to the todays outtake of 9 TWh. The potential for agricultural products as a source for biofuels is somewhat debated, which is estimated to reach as high as 39 TWh down to 7 TWh depending on source (Naturskyddsforeningen, 2014).

The outlook for using biomass as a source for transport fuel in Sweden is that there is no visible shortage to supply the entire fuel demand. The main limitation comes down to political commitment, where governmental subsidies and policies are crucial to achieve a transition towards using fuels with a renewable source (Svebio, 2008).

2.3.1 Black liquor

BL is a by-product from the Kraft pulp process (KPP) that mainly consists of water, lignin, carbohydrates from hemicellulose breakdown, cooking chemicals and other inorganic materials. These cooking chemicals is recovered and reused, typically through combustion of the BL in a recovery boiler (Ekbohm et al., 2003).

Elemental composition of BL varies depending on what type of wood is being used in the cooking process where softwood generally contains a higher content of lignin, and the operational conditions used in the pulping process (Cardoso, de Oliveira, & Passos, 2009). See Table 7 for details.

BL comes out of the cooking process generally with a moisture content of around 85 %. This is reduced in a series of flash columns called multi-stage evaporation plant, which evaporates the water content in the BL and ends up with a final dry substance (DS) of 70-80 %. At this point the BL is a thick black slurry at room temperature, with a high concentration of inorganic materials and lignocellulosic residues (Ekbohm et al., 2003).

Table 7. Elemental composition of BL (dry basis) depending on feedstock and process, weight fraction of oxygen is based on the remaining mass. Values for "softwood, pine" are based on averaged values from data (Cardoso et al., 2009).

Element	Hardwood (Kraft)	Eucalyptus (Kraft)	Straw (Soda)	Softwood, pine (Kraft)
C [%wt.]	33.2	37.3	39.1	34.85
H [%wt.]	3.3	3.6	4.5	3.45
N [%wt.]	0.08	0.09	1.0	0.065
O [%wt.]	34.52	34.91	37.97	35.585
S [%wt.]	5.2	3.4	0.8	5.15
Na [%wt.]	20.8	17.3	8.8	18.5
Cl [%wt.]	0.3	1.6	3.5	0.6
Si [%wt.]	-	-	0.23	-
K [%wt.]	2.6	1.8	4.1	1.8

2.3.2 Crude glycerol

Glycerol (C₃H₅(OH)₃) can either derive from natural sources or be synthetically produced from various processes such as soap-making or biodiesel production, where CG ends up as a by-product. In general the volumetric ratio of CG is roughly around 10 to 20 % of the total biodiesel production. Today, the increasing biodiesel production is a vast source of CG (Quispe, Coronado, & Carvalho Jr., 2013). CG is a mixture of pure glycerol and a variety of chemical and inorganic compounds where purity depends on what type of feedstock and process has been used (Hu, Luo, Wan, & Li, 2012).

As an example, the composition of some CG's can be seen in Table 8, elemental compositions of three different types of glycerol fuels depending on grade, where USP grade is pure glycerol is presented in Table 9. Methylated glycerol is the filtered sediment left after transesterification and has a purity grade of 50 to 70 %wt. glycerol, due to impurities originating from biodiesel production. This can be compared to a purity grade of 70 to 88 %wt. for demethylated glycerol, where the excess methanol used in the biodiesel production is recovered (Quispe et al., 2013). Graphical illustration of biodiesel with CG sediment is shown in Figure 8.

Besides the formation of biodiesel and the by-product glycerol, different types of organic compounds either remains unconverted in the transesterification process, for instance free fatty acids (**FFA**), or is formed, such as different kinds of soaps depending on catalyst used in the process. When the biodiesel or fatty methyl esters (**FAME**) are separated from the by-products, up to 1 % of FAME will remain left in what is called the CG. All organic components that is not glycerol will be in this case referred to as matter organic non-glycerol (**MONG**) (Quispe et al., 2013).

Table 8. Some examples on CG compositions in weight percentage (Hansen, et al., 2009).

Product	Glycerol [%wt.]	H ₂ O [%wt.]	Ash [%wt.]	Methanol [%wt.]	MONG [%wt.]
CG-1	38.4	0.3	4.2	0.14	57
CG-2	66.7	0.2	2.9	11.4	18.8
CG-3	64.5	0	3.4	13.94	18.1
CG-4	63.4	1	5.6	4.72	25.3



Figure 8. Bottle of FAME (biodiesel), with CG sediment at the bottom (eXtension, 2013).

Table 9. Example of elemental composition for glycerol at different grades (Quispe et al., 2013).

Element	USP	Methylated	Demethylated
C [%wt.]	39.1	42.05	67.27
H [%wt.]	8.7	10.14	11.43
N [%wt.]	0	<0.05	<0.05
O [%wt.]	52.2	43.32	17.06
S [%wt.]	0	0.078	<0.05
H ₂ O [%wt.]	0	1.03	1.47
Ash [%wt.]	0	3.06	2.23
Ca [ppm]		<23	119
Na [ppm]		11 600	17 500
K [ppm]		628	541
Cl [ppm]		124	154
Mg [ppm]		<8	29
P [ppm]		2 220	1 750
HHV [MJ/kg]	16.0	21.8	20.6

The main advantage of combining CG with BLG compared to burning in a CHP plant, is the ability to be able to produce more valuable commodities. Gasification instead of combustion of CG will also help reducing the problems that inorganic materials in fuel will cause when burnt, such as corrosion on gas path walls and nozzles, and slag deposits in the combustion chamber outlets and on HEX surfaces when following the gas path and subsequently cooled below melting point. CG also has a high ignition temperature (370°C) compared to for instance gasoline (280°C), which in combination with the relatively low heating value makes the fuel on its own incapable of sustaining a stable flame when combusted (Quispe et al., 2013).

2.3.3 Fermentation residues

By fermenting biomass rich of glucose and/or cellulose a variety of renewable fuels can be produced, such as ethanol and butanol. The by-product of the fermentation that is left unconverted is a lignin rich residue with an average energy content of about 18 MJ/kg (Sannigrahi & Ragauskas, 2011). As an example, a few elemental compositions of FR depending on feedstock can be seen in Table 10 and graphical representation is shown in Figure 9.

Table 10. Elemental compositions and heating values of FR (dry-ash free basis) from different feedstocks (Sannigrahi & Ragauskas, 2011).

	Hybrid poplar	Red maple	Miscanthus	Switchgrass
HHV [MJ/kg]	20.92	24.16	19.02	20.78
C [%wt.]	50.29	56.84	45.56	50.14
H [%wt.]	5.64	5.73	4.65	5.46
N [%wt.]	0.85	0.98	1.36	1.54
O [%wt.]	39.70	34.96	34.07	36.53



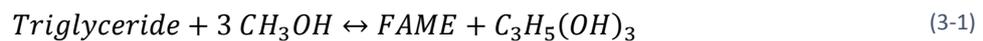
Figure 9. Photograph of FR (LAMBION Energy Solutions).

3 Theory

In this section, subjects of theoretical matter will be handled. This include equations for heating value calculation and equilibrium reactions, stoichiometric combustion of CG, and a simplified explanation of a BFP process and the transesterification process.

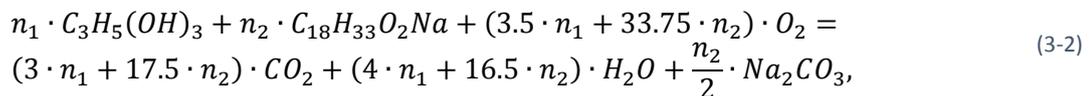
3.1 Crude glycerol

Commonly biodiesel is produced through a process called transesterification, where methanol or ethanol in combination with a catalyst, for example sodium hydroxide reacts with triglycerides to form FAME and a by-product of glycerol. The process is reversible and a simplified reaction can be seen in equation (3-1):



From biodiesel production, various amounts of by-products will end up in the CG besides glycerol, as shown in Table 8. Some of these products can be methanol, ash and water, as well as different types of MONG such as FAME, FFA, and soaps.

A stoichiometric combustion of CG containing an undetermined amount of sodium oleate (soap) with pure oxygen can be seen in equation (3-2):



where n_1 and n_2 is the amount of moles of the substance and will depend on the amount of MONG present in the CG.

3.2 Heating value

To calculate the heating value, the heat of formation from the reactants and products are used to determine the heat released from stoichiometric combustion, as seen in equation (3-3):

$$H = - \sum_{i, \text{products}=1}^n n_i \cdot \Delta H_{f,i}(T_0) + \sum_{j, \text{reactants}=1}^n n_j \cdot \Delta H_{f,j}(T_0), \quad (3-3)$$

where “i” is the products and “j” the reactants in the stoichiometric combustion. The resulting heating value gives the amount of heat that has to be removed from the combustion in order to keep temperature at reference level. The heating value acquired when mixing different fuels can be seen in equation (3-4):

$$H = \sum_{i=1}^n m_i \cdot H_i, \quad (3-4)$$

where m_i is the mass fraction of the fuel component “i” in the mixture.

While the lower heating value (**LHV**) or effective heating value is acquired by equation (3-3) and (3-4), higher heating value (**HHV**) is calculated using equation (3-5):

$$HHV = LHV + 2\,500 \cdot (8.94 \cdot m_H + M), \quad (3-5)$$

where HHV and LHV are expressed in kJ/kg, m_H is the mass ratio of hydrogen in the fuel composition, and M the moisture content of the fuel. HHV is also referred to as calorific heating value and assumes that all water components are brought back to liquid state, thus accounting for the latent heat of vaporization.

In a BLG process it is a common procedure to use sulphur-free lower heating value (**SF-LHV**), the sulphur content in the fuel is considered to leave the gasification process in reduced form with the ash melt. Calculation of SF-LHV is expressed in equation (3-6):

$$LHV_{SF} = LHV \cdot (1 - m_S), \quad (3-6)$$

where m_S is the fuel composition mass fraction of sulphur. The heating values LHV_{SF} and LHV are expressed in kJ/kg.

3.3 Equipment and labour costs

Equipment costs scale accordingly to equation (3-7):

$$C = C_0 \cdot \left(\frac{S}{S_0}\right)^k, \quad (3-7)$$

where C_0 is the reference cost of the equipment, S_0 the reference flow parameter, S the desired flow parameter, and k the scale factor.

Time where cost data are retrieved and when economic analysis is performed might be slightly different. For equipment and installation costs in the chemical industry, a common approach is to adjust the prices according to the chemical engineering price index (**CEPCI**), see equation (3-8):

$$C_{present} = C_{yearX} \cdot \frac{CEPCI_{present}}{CEPCI_{yearX}} \quad (3-8)$$

To adjust prices that increases on annual basis, such as worker wages, equation (3-9) is used:

$$C = C_{yearX} * \left(\frac{CPI_{present}}{CPI_{yearX}} \right), \quad (3-9)$$

where C_{yearX} is the cost for a specific year, CPI_{yearX} the corresponding year consumer price index (**CPI**), and $CPI_{present}$ is the CPI for the desired year. This is used to make a cost estimation for a different year than the provided data is given for.

Depreciation of the investment costs is calculated using equation (3-10) and illustrates the decreased value of the assets due to its technical lifetime expectancy from wear and tear, as well as the discount rate expected from the investment:

$$a = \frac{d \cdot (1 + d)^t}{(1 + d)^t - 1} \quad (3-10)$$

where d is the discount rate in percent, a is the annuity factor, and t is the technical life expectancy of the investment in years. The annual depreciation is then calculated using equation (3-11):

$$D = CAPEX \cdot a, \quad (3-11)$$

where D is the annual depreciation or annual capital cost of the investment in €/year, $CAPEX$ is the capital expense or total investment cost of the BFP.

3.4 Straight payback time

Time it takes for the investment to pay back in terms of revenue is calculated using a straight payback time method according to equation (3-12):

$$PBT = \frac{CAPEX}{c}, \quad (3-12)$$

where payback time (**PBT**) is expressed in number of years, $CAPEX$ is the investment cost in €, and c the cash flow of the investment in €, which is calculated using equation (3-13):

$$c_y = R_y - OPEX_y \quad (3-13)$$

where R is the revenue in € at the specific year y , $OPEX$ is the operating costs of the BFP.

3.5 Net present value

To calculate the net present value (**NPV**), which is the value of money generated over the entire lifetime of the investment if brought to present time, equation (3-14) is used:

$$NPV = \sum_{y=0}^t \frac{c_{net,y}}{(1 + r)^y} \quad (3-14)$$

where c_{net} is the net cash flow in € of the investment at the specific year y , r is the real interest rate, and t is the technical life expectancy of the investment in number of years. Net cash flow is calculated according to equation (3-15):

$$c_{net,y} = (R_y - (OPEX_y - Labour\ costs) - D) \cdot (1 - Tax) + Labour\ costs + D, \quad (3-15)$$

where R is the revenue in € at the specific year y , $OPEX$ is the operating costs of the BFP, Tax is the income tax for corporations, and D the calculated annual depreciation cost.

3.6 Internal rate of return

Internal rate of return (**IRR**) is the rate of which the invested capital will return over the course of the expected technical lifetime, this is further described in equation (3-16):

$$0 = CAPEX - \sum_{y=1}^t \frac{c_{net,y}}{(1+i)^y}, \quad (3-16)$$

where i is the IRR, $c_{net,y}$ is the net cash flow at the specific year y , t the technical lifetime expectancy, and $CAPEX$ the capital investment of the BFP. The value of i is iterated so that the sum of the equation ends up in zero.

4 Method

In the following section the methodology used in the thesis is described, such as parameters used in simulations and costs for different material and energy streams.

4.1 Aspen® simulation model

A model to determine the output of different material and energy flows was made in a computer software from AspenTech called Aspen Plus® V8.4, and was restricted to the process after the EFG. After the model was constructed, the input flow and composition of the SG was modified for each case. With design specifications for steam pressure, temperature, and flow, the idea of the concept was to determine important material and energy streams while maintaining the necessary amount of steam to sustain the PM as well as the BFP itself. The system boundaries were set according to Figure 10. The net power output from the BFP process was calculated by creating and sum up all work streams from pumps, compressors, and turbines. The amount of tail gas that was assumed taken out of the MSY loop was set to 4 % according to previous model built (Andersson, Lundgren, & Marklund, 2014) except for case 7, where this would exceed the amount of fuel gas needed for the lime kiln. Instead, the value for the tail gas split was iterated and finally set to 2 % of the recycled gas flow. Do note that in practice tail gas bleed flow is never regulated to satisfy lime kiln fuel supply, but to prevent build-up of inert components in the recycled gas flow.

The composition and mass flow of the input SG stream is calculated based on simulations on the BLG, which is further described in section 4.1.1. For each case, all material and energy streams were documented for later assessment in the techno-economic analysis.

Due to confidential information on the acid gas removal (**AGR**) unit was constructed as a “black box” where output and input streams from this unit were modified according to known parameters. The unit modelled was a Rectisol™.

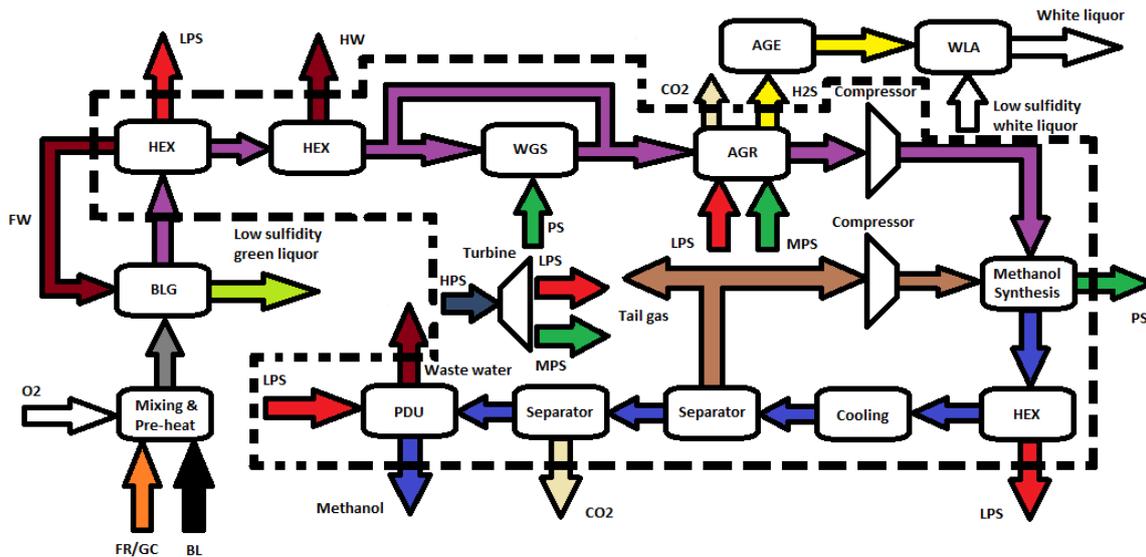


Figure 10. Illustration of system boundary of Aspen® model with material streams, where the dashed line represent the boundary. Arrows covered and exiting the boundaries are calculated but not further handled, while the arrow entering have a fixed input.

The Aspen® model designed for this thesis did not handle the resulting output streams from the AGR other than the SG flow, for instance the stream to the acid gas enrichment (**AGE**) or the white liquor adsorption (**WLA**). The equipment handling these streams was included in the overall BFP investment cost.

Due to the complexity and/or lack of data for different components, all temperature and pressure losses in piping, HEX's, heaters, and reactors were neglected. The WGS and MSY was modelled as a hierarchy to minimize the amount of blocks and components in the main flow sheet. For a detailed description of settings for blocks and streams, see Appendix 1 – Aspen® modelling settings.

4.1.1 Synthesis gas composition

In order to estimate the composition of the SG input from gasification (see section 1.4) into the MSY model with either pure BL or BL and CG/FR mixture, a simulation was made using a Matlab scripted program called SIMGAS. Calculation of the different blend compositions were made in Excel to determine the feedstock composition, for a detailed view of calculation procedure see Appendix 2 – Feedstock composition calculation sheet. All input parameters were assumed constant.

SIMGAS works under the assumption that ideal mixtures are present and with a fixed fraction of methane, which is based on empirical figures for BLG with Chemrec's EFG. The program uses the basis of thermodynamic equilibrium and minimization of Gibbs energy, which is a non-linear problem and the input element balances are used as a linear constraint to approximate a linear problem. The program uses a set of chemical components that are supposed to be present in the equilibrium mixture, and the quantity is calculated based on a minimization algorithm that uses thermodynamic data and element balance of the components. Oxygen supplied in the process was determined based on iteration from a specified oxygen concentration depending on the fuel input (Carlsson, Marklund, Furusjö, Wiinikka, & Gebart, 2010). For calculations performed in this thesis, conditions were defined according to Table 11, concentration of H₂S was determined according to an empirically modified thermodynamic model based on BLG.

Table 11. SIMGAS settings for simulation of SG elemental composition, temperature, and mass flow.

Parameter	Setting
Atomizer gas (oxygen 99.5 %vol.)	Supplied at 100°C, 35 bar
CH ₄ cold gas volume fraction	1 %
EFG effect loss	500 kW/unit
EFG reactor pressure	31 bar
EFG reactor temperature	1273.15 K
Primary/secondary quench spray	200°C, 35 bar

Black liquor composition

The composition of the BL from Vallvik PM used for SG composition simulation is described in detail in Table 12. Other properties for BL for simulation of SG composition were defined as presented in Table 13. All these values was assumed constant.

Table 12. Elemental composition of BL used for simulation of SG composition (dry basis) (Andersson, Lundgren, & Furusjö, 2013).

	[%wt.]
C	33.6
H	3.4
S	4.7
O	35.1
Na	21.4
K	1.3
Cl	0.6
N	0.0

Table 13. Defined variables for BL for SG composition simulations, values of HHV and \dot{m} in DS (Andersson et al., 2013).

Variable	Value	Unit
\dot{m}	1 500	ton/day DS
HHV	13.56	MJ/kg DS
DS	73.2	%

Crude glycerol composition

Due to the varying composition of a CG, pre-determined variables were set for medium blend quantity as described in Table 14. For medium blend quantity, different properties on the CG was used to determine if there might be any possible influence on the composition of the SG by varying the quality of the CG. The default amount of MONG contained in the CG was set to 25 %wt.

Table 14. Crude glycerol composition at three different setting, units in weight percentage and DS. For simplification in composition calculations, MONG was defined as sodium oleate (C₁₈H₃₃O₂Na).

	Glycerol/MONG [%wt.]
Crude glycerol	50/50; 75/25; 100/0

To simplify the calculations for the CG composition, assumption where made that the CG only contained pure glycerol and MONG. All the other components that can be found in CG, such as methanol is assumed to be either negligible or removed before mixing with BL. The minor presence of ash in the CG is also assumed to have no significant effect on the performance of the process as a whole, which is why this was also neglected. The elemental composition for CG used in the simulations is presented in Table 15 and Table 16.

Table 15. Elemental composition (dry basis) for all CG quantities used in simulation of SG composition, with default settings for MONG.

	[%wt.]
C	47.10
H	9.30
S	0
O	41.71
Na	1.89
K	0
Cl	0
N	0

Table 16. Elemental composition (dry basis) for CG at medium blend used in simulation of SG composition, at different CG quality than default.

	MONG 0 % [%wt.]	MONG 50 % [%wt.]
C	39.12	55.07
H	8.76	9.84
S	0	0
O	52.12	31.31
Na	0	3.78
K	0	0
Cl	0	0
N	0	0

The DS for the CG was set according to (Hansen, et al., 2009) and its average values for moisture content in various CG's. This gave the CG a DS value of 94.5 %.

Fermentation residue composition

The elemental composition and heating value of FR from fermentation of hybrid poplar were used in the calculations (see Table 10), due to its closest relation to the desired type of feedstock for the cases. The data from Table 10 were missing 3.52 % to end up in total weight percentage of 100 %, this difference was distributed along each element according to their weight ratio.

According to Sannigrahi & Ragauskas (2011) the amount of potassium in hybrid poplar is around 25 000 mg/kg of FR, this figure was added to the elemental composition which was further adjusted to satisfy 100 % in weight percentage. The resulting elemental composition can be seen in Table 17.

Table 17. Elemental composition (dry basis) of all FR quantities used for simulation of SG composition.

	[%wt.]
C	52.11
H	5.84
S	0
O	41.145
Na	0
K	0.025
Cl	0
N	0.88

The amount of DS for the FR was assumed to be the same as BL (73.2 %). According to Furusjö (2015) who received figures from Inbicon A/S, the DS in their samples ends up in around 50 % after filtration of the FR. This might cause problem such as low temperatures in the gasifier, which is why pre-treatment of the samples was assumed to be performed, preferably drying through waste heat utilization.

Heating values

The HHV for BL in Table 13 (13.56 MJ/kg DS) was used in all cases to calculate the heating value of the fuel supplied to the EFG. In cases involving fuel mixtures of BL and FR/CG equation (3-4) was used to determine the HHV of the mixture.

By using equations (3-2) and (3-3) with the heat of formation values presented in Table 18, the HHV for CG was calculated for different CG qualities, which is presented in Table 19. All calculations were assumed to be at reference temperature level of 25°C.

Table 18. Heat of formation for products and reactants in stoichiometric combustion of CG at reference temperature (The National Institute of Standards and Technology, 2011).

Component	Stoichiometric property	ΔH_f^0 [kJ/mol]
CO ₂	Product	-393.52
H ₂ O	Product	-241.83
Na ₂ CO ₃	Product	-1130.77
C ₃ H ₅ (OH) ₃	Reactant	-669.6
C ₁₈ H ₃₃ O ₂ Na	Reactant	-764.8
O ₂	Reactant	0

Table 19. Heating values of CG depending on quality. Default HHV value is used in all cases involving CG, and the 0 and 50 % MONG values are only used for evaluation of the possible influence different quality on CG might have on the methanol production.

	MONG 0 %	MONG 25 % (default)	MONG 50 %
HHV [MJ/kg DS]	18.01	22.89	27.76

For FR the HHV was taken from Table 10 for hybrid poplar, which is 20.42 MJ/kg DS. All the presented values in this section were used as input for simulation of SG composition depending on fuel composition.

Heating value of tail gas was calculated using heat of formation of tail gas with stoichiometric combustion of components as previously described, with simulation data from case 1; mass flow of tail gas of 0.294 kg/s and a mass fraction of hydrogen of 15.0 %. Minor components such as methanol and hydrogen sulphide were neglected. Using data from Table 20 and Table 21 gave tail gas a HHV of 37.61 MJ/kg, which was assumed to be constant for each case.

Table 20. Tail gas composition output in case 1 from simulations, pure BL as fuel.

Component	Mole flow [kmol/hour]
CO	4.391
H ₂	78.62
CH ₄	19.91

Table 21. Heat of formation values for tail gas components used for calculating heating value of tail gas (The National Institute of Standards and Technology, 2011).

Component	Stoichiometric property	ΔH_f^0 [kJ/mol]
CO	Reactant	-110.53
H ₂	Reactant	0
CH ₄	Reactant	-74.87
O ₂	Reactant	0
CO ₂	Product	-393.52
H ₂ O	Product	-241.83

4.2 Techno-economic analysis

In order to evaluate the performance of a BFP integrated in the Rottneros Vallvik PM, a techno-economic analysis was performed. Besides this, net efficiency of the BFP was also evaluated to gain a weighted perspective on which case was likely the favourable one. Tax was calculated at 22 % of net income according to the income tax for corporations in Sweden.

4.2.1 Material and energy flow costs and efficiency

According to Andersson et al (2013), the availability of BL for the case study of Rottneros Vallvik PM is 90 % in 348 days on annual basis. This gave a total of 7 517 annual operating hours, which were used in combination with material and energy costs and the simulated flows of all energy and material streams to determine the annual operational costs and revenues.

Each material and energy stream is sorted either as input or output, where the input streams are added to the operating expenses (**OPEX**) and the output streams to the revenue of the BFP. Full capacity for methanol production of the plant was considered to be achieved not until after year three due to plant construction. Assumptions were made that for the first operational year the methanol production would be at 60 % of full capacity, for the second, third, and fourth year it would reach up to 80, 95, and 100 % respectively (Andersson et al., 2013).

Costs and energy efficiency for the different material and energy streams used for calculations is presented in Table 22. Energy was converted according to their respective electricity equivalents and total plant efficiency was calculated based on the difference in material and energy streams compared to a reference case where a recovery boiler was used. Compared to the reference case there was an increased use of lime kiln fuel by 25 % which is deducted from the tail gas production used for lime kiln fuel replacement. Power and steam production was no longer linked to the recovery boiler but instead produced in a separate boiler and in the BFP processes. This increased electric power import was compared to the import for the reference case. Even though an ASU was not considered in terms of cost installation and electric power usage, the power to produce the oxygen supplied to the BFP was still considered. Figures used for calculation of amount of electric power used was 1.43 MW per kg/s of oxygen supplied to the BFP process.

Table 22. Price data and electric conversion efficiency for material and energy streams used in techno-economic analysis (Andersson et al., 2013).

	η_{el} [%]	Price	Unit	Type
Bark/logging residues	46.2	19.3 ¹⁾	[€/MWh]	Input
CG	46.2	20 ²⁾	[€/MWh]	Input
Electricity	100	45.8 ³⁾	[€/MWh]	Input
FR	46.2	40 ⁴⁾	[€/MWh]	Input
Methanol	55.9	See Table 23	[€/tonne]	Output
O₂	-	69	[€/tonne]	Input
Tail gas	57.6	46	[€/MWh]	Output

1) (Swedish Energy Agency, 2015), 2) (Quispe et al., 2013),

3) (SCB - Swedish Statistics, 2008), 4) (Furusjö, 2015).

According to Swedish Energy Agency (2015) there is a substantial fluctuation in the ethanol price, which was used as a selling price assumption for methanol due to the similar nature of the fuel. With an ethanol price that ranged from 560 to 880 € per tonne, the lower, upper, and the average of 720 €/tonne were used to assess the BFP profitability. The quota of energy content in methanol (LHV of 20.1 MJ/kg) and ethanol (LHV of 25.8 MJ/kg) was used to convert the price for methanol due to lower energy content, as presented in Table 23. No methanol recovery from CG has been considered or taken into account. The default price used for calculation of methanol revenue was the average price of 560.8 €/tonne or 100.5 €/MWh.

Methanol from fossil origin, usually from natural gas, has according to Methanex Corporation (2015) an average European pricing in 2015 of 381.0 €/tonne or 68.3 €/MWh (Methanex Corporation, 2015). This was used to determine the profitability of the BFP when performing a price matching between methanol from fossil and renewable sources.

Table 23. Different price ranges for methanol used for techno-economic analysis of BFP.

	Fossil	Minimum	Average	Maximum
Methanol price [€/tonne]	381.0	436.1	560.8	685.4
Methanol price [€/MWh]	68.3	78.1	100.5	122.8

The price for bark/logging residues was based on figures from 2014 by (Swedish Energy Agency, 2015) for solid by-products from industries. Currency in the report was converted into EUR. The flow of bark/logging residues into the boiler was calculated based on the heating value of the fuel, the amount of HP steam necessary in the BFP process, and some boiler parameters. See Appendix 3 – Boiler calculations to determine biomass flow for details. This generated a ratio for each MW of steam produced to MW of fuel input into boiler of 90.7 %.

The price for CG varies depending on quality and source of purchase. Assumptions on quality and thereby price were made and adjusted with some margins. According to Quispe et al (2013) there was an average price on CG of 10 cents per kilogram in 2012, which would correspond to roughly 17.5 €/MWh with assumed CG composition, while figures from Gholami et al (2014) showed an average CG price in 2011 of 8 cents per pound, or 23.6 €/MWh (Gholami, Abdullah, & Lee, 2014). Price for pure glycerine in Europe (2012) ranged on average as high as 694.3 €/tonne (119.3 €/MWh) (Quispe et al., 2013).

For electricity the price was determined using figures from January 2007 by SCB - Swedish Statistics (2008) and the price for the largest industrial energy consumer “li” with an annual energy consumption of 70 000 MWh and a maximum power demand of 10 000 kW. Prices were converted using the individual CPI for electricity (average value in 2007 of 129.1, in 2014 to 2015 of 104.6) and equation (3-9) to estimate the current price.

HW was assumed to be either used internally as drying of feedstock or used in various processes that did not require high temperatures of source of heat. Any heat not utilized in the BFP or PM was assumed to be cooled away as waste heat.

The cost for the oxygen supplied to the BLG process was assumed to be purchased “over the fence”, thus the cost for electricity usage that would have been required for an air separation unit (ASU) was not accounted for. The electric power used in the ASU to produce oxygen was calculated to 1.43 MW per kg/s, based on figures from an ASU model (Kooijman, 2006). With the same efficiency for oxygen as for electricity, this was used in the calculations of the total electric efficiency of the BFP process.

Assumptions were made that the tail gas originating from the MSY process would be used to replace fuel used in the lime kiln, thus the price for the tail gas was based on to the reduced cost of import of lime kiln fuel. According to Ekbohm et al (2003), the fuel demand for the lime kiln will increase by approximately 25 % or 4.5 MW when processing BL in an EFG instead of a recovery boiler, which was accounted for in the calculations.

An annual price increase on all material and energy streams were calculated using the CPI presented in Table 29 and added to the OPEX and revenue in the techno-economic analysis.

The energy efficiency calculation process that was performed is illustrated in equation (4-1), where all streams are expressed in terms of MW after electric equivalent conversion, heating value for all materials in SF-LHV.

$$Energy\ efficiency = \frac{Methanol_{output}}{Bark + logging\ residues + Oxygen_{el.} + ((Lime\ kiln\ fuel) \cdot 0.25 - Tail\ gas)} + CG\ or\ FR + (Electricity_{input} - Electricity_{reference})} \quad (4-1)$$

4.2.2 Equipment and labour costs

The capital expense (**CAPEX**) is the total direct and indirect investment costs that was estimated to determine how well an investment in the project BFP would end up. All costs were based on data from Stare et al (2012) which is an in-house report. Due to confidential material, all costs for equipment were calculated individually but presented as a total investment cost. All investment costs were calculated using equation (3-7) except for the EFG, where the rounded up number of “trains” was decided by the desired input effect of the fuel and the maximum fuel feed capacity for each unit (Andersson et al., 2013). Scaling factor and parameter, and reference capacity for each unit that was calculated is shown in Table 24.

Costs in CAPEX not related to equipment, but to changes on-site such as re-routing of roads, ground works, avoided costs, and piping to name a few were included. These are fixed costs and therefore not affected by any change in plant size. Like previously described, no cost detail of this will be provided due to confidential information, instead this post was included in the total CAPEX.

Calculations of depreciation were made after calculation of CAPEX with parameters shown in Table 25 using equations (3-10) and (3-11).

Table 24. Scaling factor and reference capacity for different units in the BFP process used for CAPEX calculation.

Unit	Reference capacity	Scaling factor	Scaling parameter
EFG	70 MW feedstock LHV*	-	Number of trains
WGS	26 329 kg/hour	0.65	Dry gas flow
AGR	831 kmol/hour	0.65	CO ₂ -flow
MSY	17 838 kg/hour	0.65	Methanol flow
PDU	17 838 kg/hour	0.65	Methanol flow
AGE	-	-	Same for all
WLA	-	-	Same for all
Methanol storage and handling	17 838 kg/hour	0.65	Methanol flow
BOP	17 838 kg/hour	0.65	Methanol flow
Thermal oxidizer	-	-	Same for all
Flare stack with safety zone	17 838 kg/hour	0.65	Methanol flow

*For cases with medium and high blend of CG or FR, reference capacity is 85 MW of feedstock LHV.

Besides the required amount of “trains” necessary for the fuel feeding capacity, one redundant EFG in case 1 for utilization under maintenance was accounted for in the calculation for CAPEX. No redundant units were used in calculations for CAPEX for any case involving additional fuel feedstock, one unit down would still be sufficient for gasifying pure BL and assumptions were made that methanol production would instead be lowered during downtime to just using BL as fuel. The price of the EFG includes fuel handling as well as any pre-treatment necessary prior to gasification.

Due to lower ash contents in additional feedstock thus lowering the melt concentration of the fuel mixture, assumption were made that the feedstock load capacity of the BLG-unit would increase to 85 MW when subjected to medium or high blend quantity of CG or FR.

Table 25. Parameters for calculation of depreciation.

Parameter	Value	Unit
Technical life expectancy	25	Years
Discount rate	8	%

CAPEX costs were adjusted using equation (3-8), according to CEPCI indices from Table 26.

Table 26. CEPCI index for CAPEX data (Oct. 2012) and latest available (Oct. 2014).

Year	CEPCI
2012	584.6
2014	579.8

The labour costs are part of the OPEX and were made using some assumptions. For starters, the amount of employees and their salaries was based on numbers from Philips et al (2007) with some modifications:

- Yard employees were kept at the original number of 32 instead of 12
- Clerks and secretaries were kept at the original number of 5 instead of 3
- Salaries (see Table 27) from the source were used and recalculated from USD to EUR from currency trade rates, then raised according to CPI to 2014 levels using equation (3-9) and data from Table 29
- A payroll tax of 31.42 % (Ekonomifakta.se, 2015) was added to the total operating cost to cover for costs such as worker insurances, healthcare, and retirement funding

Table 27. Salaries and number of employees at different positions used for calculation of operating costs (Philips, Aden, Jechura, Dayton, & Eggeman, 2007).

Position	Number	Salary 2002 [\$]	Salary 2002 [€]
Plant manager	1	110 000	116 742
Plant engineer	1	65 000	68 984
Maintenance supervisor	1	60 000	63 678
Lab manager	1	50 000	53 065
Shift supervisor	5	45 000	47 758
Lab technician	2	35 000	37 145
Maintenance technician	8	40 000	42 452
Shift operators	20	40 000	42 452
Yard employees	32	25 000	26 532
Clerks and secretaries	5	25 000	26 532

The number of employees were modified so that there was an increase in number of shift operators, maintenance technicians, and yard employees. The assumed number of increased staff was based on the feedstock heating value and is presented in Table 28.

Table 28. Increase in number of employees depending on case.

Case	2	3	4	5	6	7
Shift operators	+2	+2	+2	+2	+2	+4
Maintenance technicians	+2	+2	+2	+2	+2	+4
Yard employees	+1	+1	+1	+1	+1	+2

The CPI is a method to measure price development and is often used to determine rate of inflation, taxing, product costs, and income increase. Figures used for calculation of present salaries can be seen Table 29.

Table 29. CPI average values in Sweden for specific years, where index=100 is from 1980 (SCB - Statistics Sweden, 2015).

Year	CPI
2002	272.8
2014	313.49

An annual cost increase on salaries was calculated using the CPI presented in Table 29 and added to the OPEX in the techno-economic analysis.

4.2.3 Summary

To perform a techno-economic analysis, calculations of PBT, NPV, and IRR was performed using equation (3-12), (3-14), (3-15), and (3-16). NPV was calculated with a real interest rate of 4 %. PBT was iterated so that the sum of cash flows was consistent with the CAPEX.

4.2.4 Sensitivity analysis

A sensitivity analysis was performed in order to determine the influence each revenue and cost had on the outcome. The change of costs were performed at $\pm 30\%$ of the default cost and plotted against the change in IRR and NPV of the investment. The greater the slope of the line, the greater a change in cost of that specific post will affect the profitability of the investment. An analysis to estimate the sensitivity to revenues, OPEX, and CAPEX was performed. This was followed with another analysis on each post to determine which one had the greatest influence on the BFP profitability

5 Results and analysis

The following section will present the results acquired from modelling and calculations performed in this thesis, which will be followed by a brief analysis.

5.1 Aspen® simulation model

From the system boundaries represented in Figure 10, the resulting model for calculation of input and output streams can be seen in Figure 11. By changing the input SG (SG from EFG) composition and flow, data for every case is calculated. The two hierarchies of the WGS and MSY can be seen in Figure 12 and Figure 13.

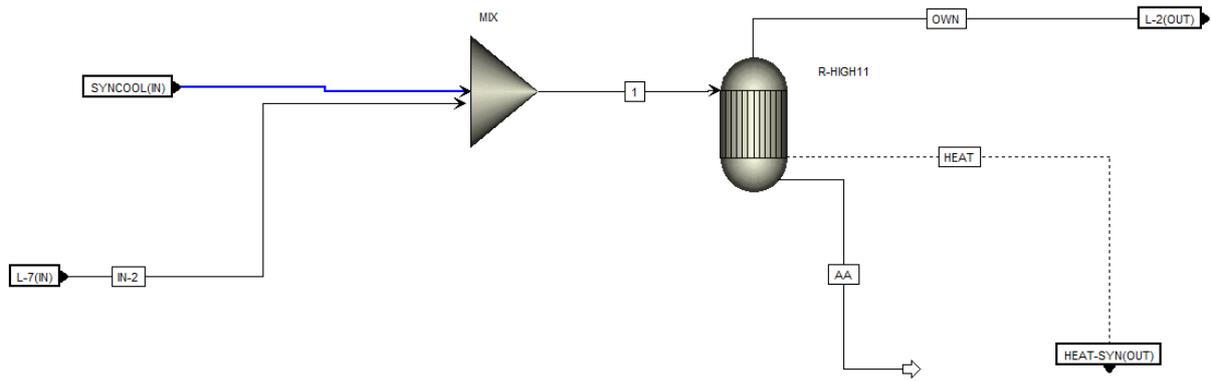


Figure 13. Sub-model (hierarchy) of the MSY process in Aspen® model, main block shown in Figure 11.

The results from simulations with SIMGAS and the Aspen® model for all cases are shown in Table 30.

Table 30. Steam, energy, and material flows after gasification for different cases, values from simulation of SIMGAS and Aspen® model. All material flows are with SF-LHV. HW effect is based on an output temperature of 95°C with a ΔT of 85°C. The electric power is the increased import power compared to the standard case of having a recovery boiler and the loss of power production.

Case	Fuel input [MW]	O ₂ [ton/h]	Dry SG [MW]	HW [MW]	Steam HP [MW]	Power [MW]	Tail gas [MW]	Methanol [MW]
1	196.1	22.0	132.2	5.8	83.4	18.3	10.1	99.6
2	232.6	25.3	162.8	7.0	79.5	20.1	12.7	122.7
3	278.3	29.2	201.4	8.5	74.5	23.5	16.0	151.6
4	337.0	34.2	251.0	10.4	66.8	27.1	20.4	188.9
5	235.7	25.2	166.5	6.9	80.2	20.1	12.2	125.7
6	349.2	34.6	265.9	10.4	68.4	28.1	18.4	201.4
7	553.7	51.4	444.4	16.5	53.3	42.5	20.3	341.9

An overview of the results show an increase of methanol output by 23 to 26 % by adding 10 %wt. of additional fuel feedstock, which is further increased at higher input. At blend rate of 50 %wt. of CG which is an increase of 100 % of total dry mass flow, the methanol output has increased by nearly 243 %.

The energy efficiency for each case, when comparing to a reference case using a recovery boiler to produce heat and power, is presented in Figure 14.

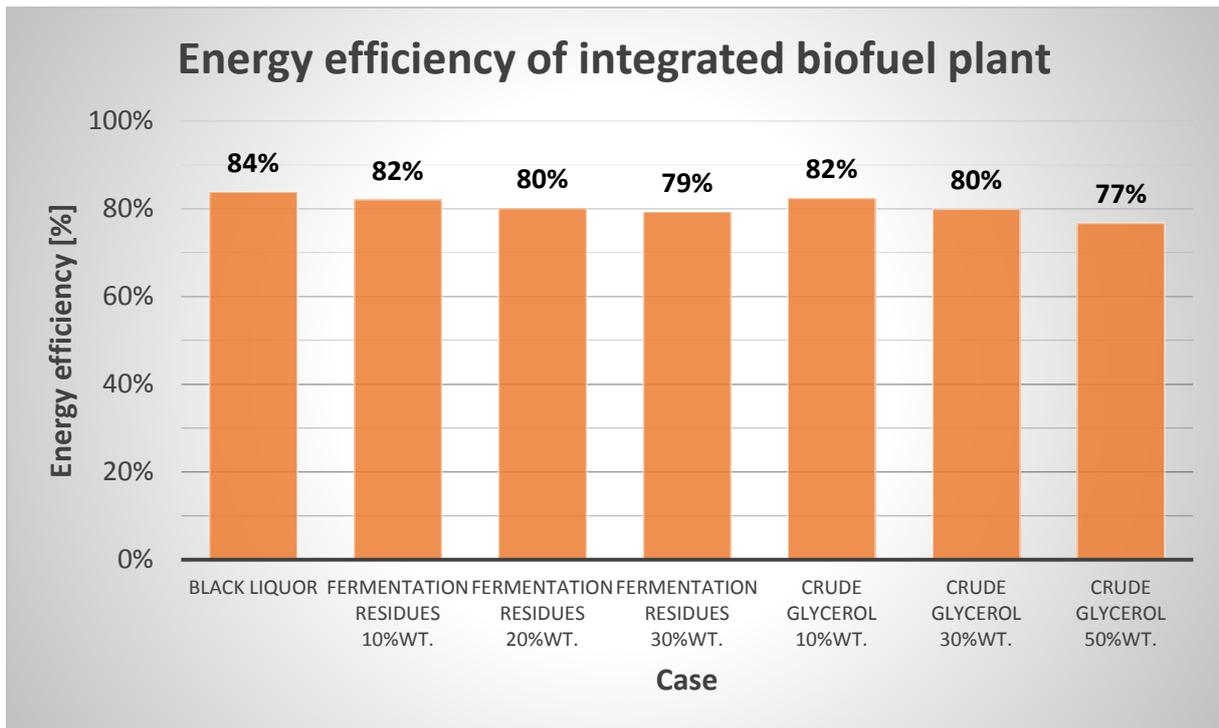


Figure 14. Electric efficiency of methanol production for integrated BFP, when compared to reference case parameters.

A quick survey of the resulting change in the energy efficiency for each case shows a modest decline at higher blend rates, with a maximum drop of 7 %-units at 50 %wt. of CG. Higher input of additional feedstock, in combination of a lower electric power production and increased use of electric power in the BFP processes is the main cause why higher blend rates has a lower energy efficiency than just using BL as feedstock for methanol production.

One important parameter worth looking into is the conversion efficiency, which is the quota of the feedstock input and the output methanol energy in terms of sulphur-free LHV. The sulphur-free LHV is used to better show the conversion efficiency of the organic part of the feedstock, since the inorganic goes back to the PM as white liquor. This gives an indication on how well the fuel converts through the process, and is dependent on operating parameters as well as input fuel composition. In Figure 15 the conversion efficiency is shown depending on which case is of interest. Note that there is no change in operating parameters, just fuel composition input.

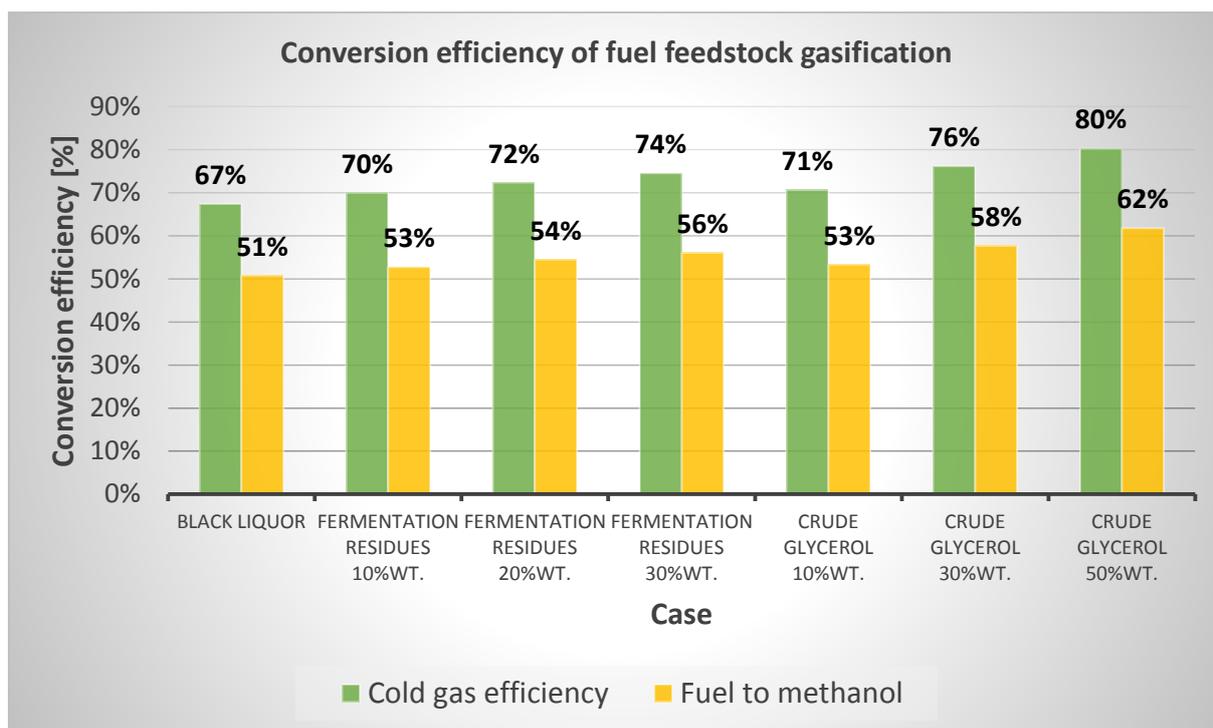


Figure 15. Cold gas efficiency and conversion efficiency of fuel in EFG-unit to methanol, depending on case setup. Units based on sulphur-free LHV.

The figure above shows a detectable improvement in the conversion efficiency with increasing additional fuel input, this is due to the lower concentration of inorganic components in the fuel mixture as well as higher concentration of carbon and hydrogen of the input fuel which will generate an increased heating value. At blend ratios exceeding 30 %wt. DS the fuel conversion efficiency has increased from the BL case of 51 % with 5-7 %-units, the higher figure is for BL and CG mixture. For the CG high blend (case 7) there is an increase of 11 %-units in methanol conversion efficiency compared to the case of pure BL. Further improvement for the conversion efficiency of fuel to methanol could have been performed by simply adjusting the “bleed” of tail gas. Lowering the bleed would have yielded a higher CAPEX due larger methanol flows and a reduced flow of replacement gas for the lime kiln fuel, but a higher methanol output.

For comparative purposes, any influence of a different quality of the CG is evaluated in the same procedure. This is performed in order to detect if the change in CG composition has any noticeable influence on the process after the BLG. Results of the comparison are presented in Table 31, with a comparative figure from the default setting.

Table 31. Comparative view of steam, energy, and material flows after BLG with different CG quality for a medium blend quantity, with red figure representing the default setting of 25 %wt. MONG. All material flows are with SF-LHV. HW effect is based on an output temperature of 95°C with a ΔT of 85°C. The electric power is the increased import power compared to the standard case of having a recovery boiler and the loss of power production.

Case	Fuel input [MW]	O ₂ [ton/h]	Dry SG [MW]	HW [MW]	Steam HP [MW]	Power [MW]	Tail gas [MW]	Methanol [MW]
CG medium	313.8	32.3	232.4	9.6	67.1	26.8	16.6	175.8
0 % MONG	(349.2)	(34.6)	(265.9)	(10.4)	(68.4)	(28.1)	(18.4)	(201.4)
CG medium	384.6	37.0	299.3	11.3	70.4	29.5	20.3	226.9
50 % MONG	(349.2)	(34.6)	(265.9)	(10.4)	(68.4)	(28.1)	(18.4)	(201.4)

As with the project cases, the conversion efficiency is calculated and compared depending on the MONG concentration where 25 %wt. is the set default value. The results from figures in Table 31 is illustrated in Figure 16, and shows a slight change in conversion rate depending on the MONG concentration in favour of higher figures. Sodium oleate have a higher carbon and hydrogen concentration than glycerol, which in turn will generate a SG with higher concentrations of carbon monoxide and hydrogen. This gives the impression that larger amount of impurities, excluding water, is favourable compared to pure glycerol.

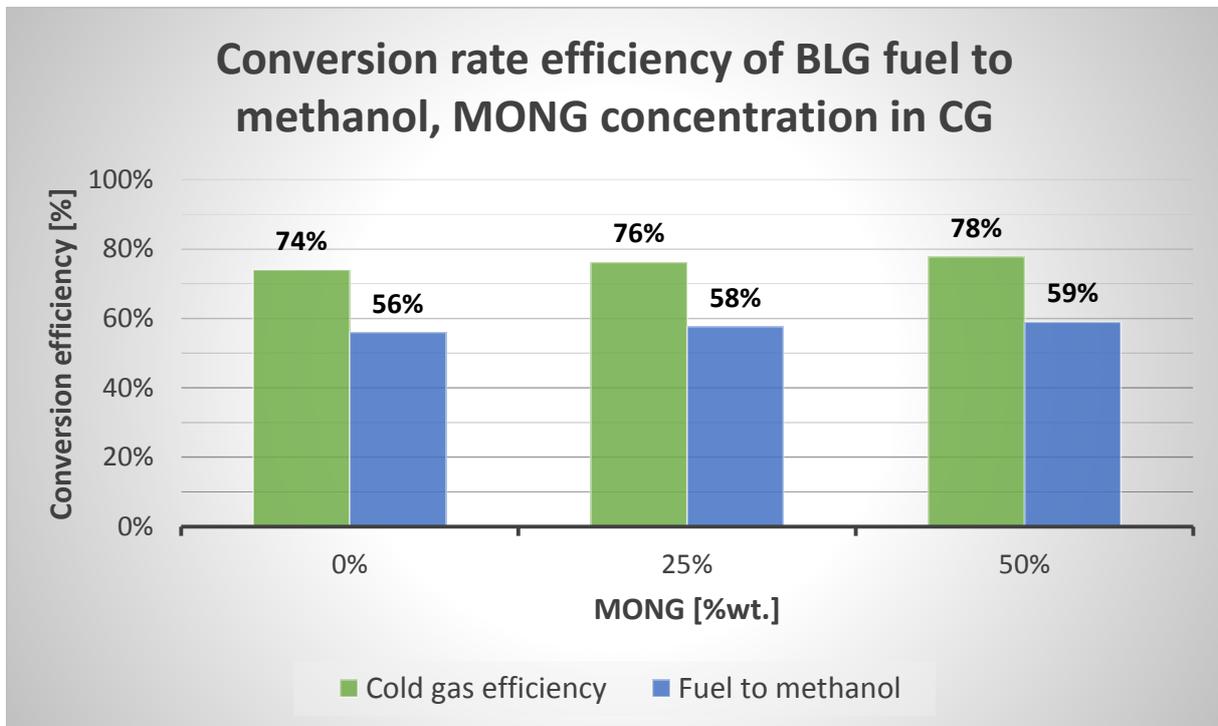


Figure 16. Change in methanol conversion efficiency depending on MONG concentration in CG.

From the results in Table 31 there is a detectable change in the input/output streams depending on the variation of CG quality. Higher concentration of MONG leads to an increased SG output and thereby an increase in methanol production, which in terms leads to a higher power demand to compressors for pressure increase in the MSY process.

5.1.1 Synthesis gas composition

From the method previously described in section 4.1.1, the result of the simulation of SG composition for all cases is presented in Figure 17, which is used in the Aspen® model for evaluating steam, material, and energy flows after the EFG. These flows are previously presented in Table 30.

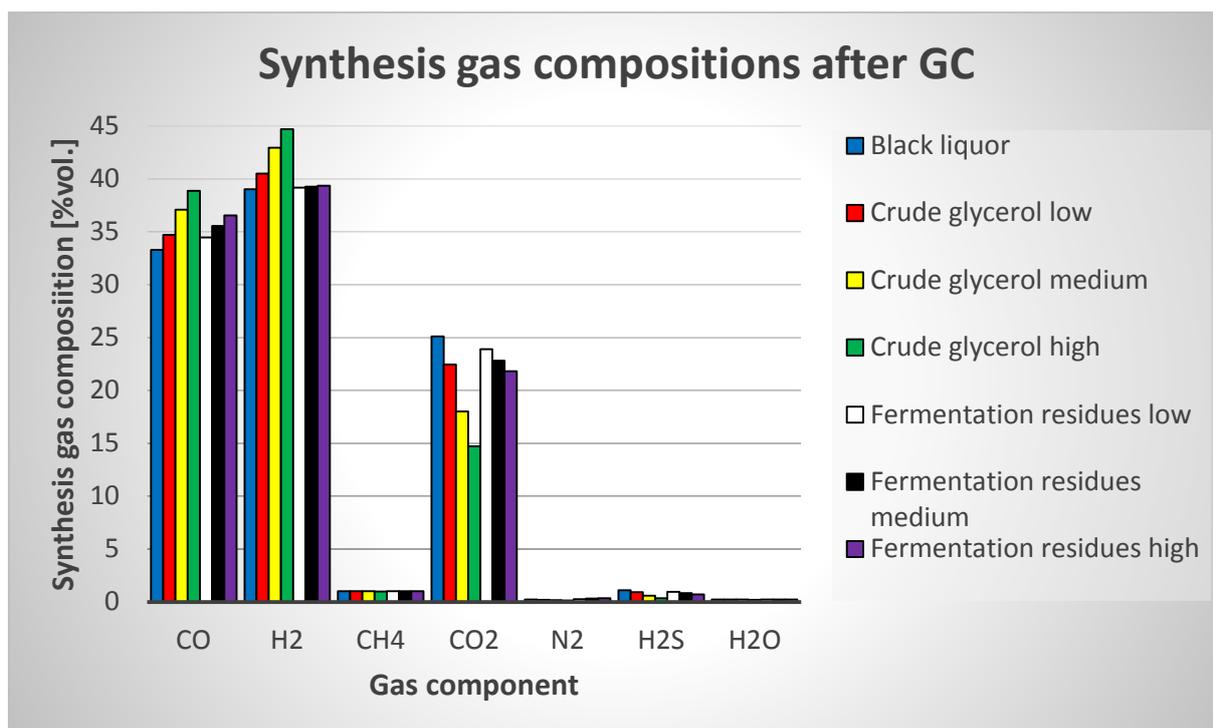


Figure 17. Composition of SG from simulations for different case setups, units in volume percentage. Conditions are after gas cooling (GC) and removal of excess water content.

Compared to the reference case where only BL is involved, concentrations of carbon monoxide is increased by approximately 1.2 %vol. for each 10 %wt. DS when adding CG or FR to BLG, with slightly higher figures for CG than FR. Concentration of hydrogen is only increasing at increased blend rates for CG while for FR it remains almost constant. One noticeable difference is also that CO₂ concentrations is dramatically reduced for CG while there is a smaller drop in the blend cases with FR. Concentrations of hydrogen sulphide is also dropped for both CG and FR at higher blend rates, where the largest drop occurs for CG.

For comparative purposes, the influence of a different quality of the CG is evaluated in the same procedure to detect any change in the SG composition. The resulting SG composition is presented in Table 32 and compared with the default setting for CG with a MONG content of 25 %wt.

Table 32. SG composition depending on MONG concentration, where the column furthest to the right is for the default value for MONG concentration of 25 %wt.

Element	CG medium 0% MONG [%vol.]	CG medium 50% MONG [%vol.]	CG medium [%vol.]
CO	35.08	38.84	37.08
H ₂	41.54	44.05	42.94
CH ₄	1.00	1.00	1.00
CO ₂	21.24	15.28	18.02
N ₂	0.17	0.16	0.17
H ₂ S	0.74	0.45	0.58
H ₂ O	0.22	0.21	0.22

As seen in Table 32 there is a noticeable shift in the composition in the SG, mainly for CO and H₂ at the expense of CO₂ and H₂S. The change in SG composition due to change in CG quality is graphically illustrated in Figure 18 and Figure 19. Do note that results are from simulations from SIMGAS which makes assumptions previously described in section 4.1.1. Since no pilot testing has been performed to confirm the resulting gas output composition, the acquired results cannot be considered accurate.

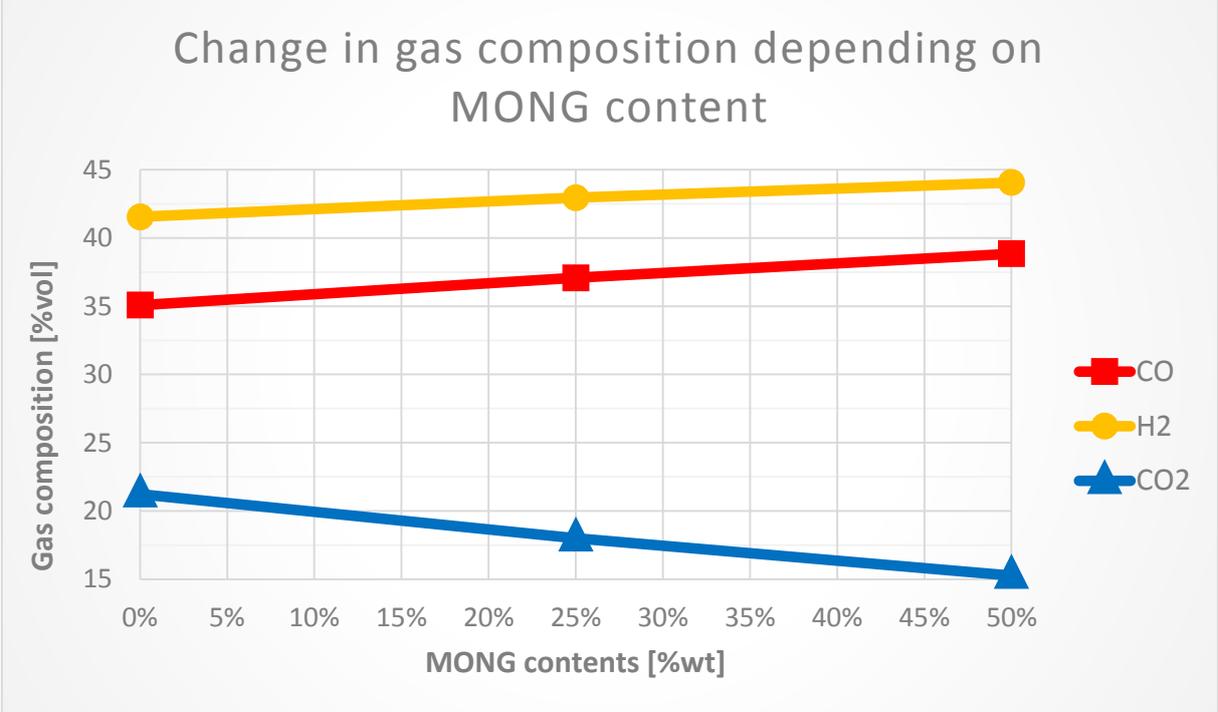


Figure 18. Illustration of change in SG composition for CO, H₂, and CO₂ depending on CG quality and content of MONG.

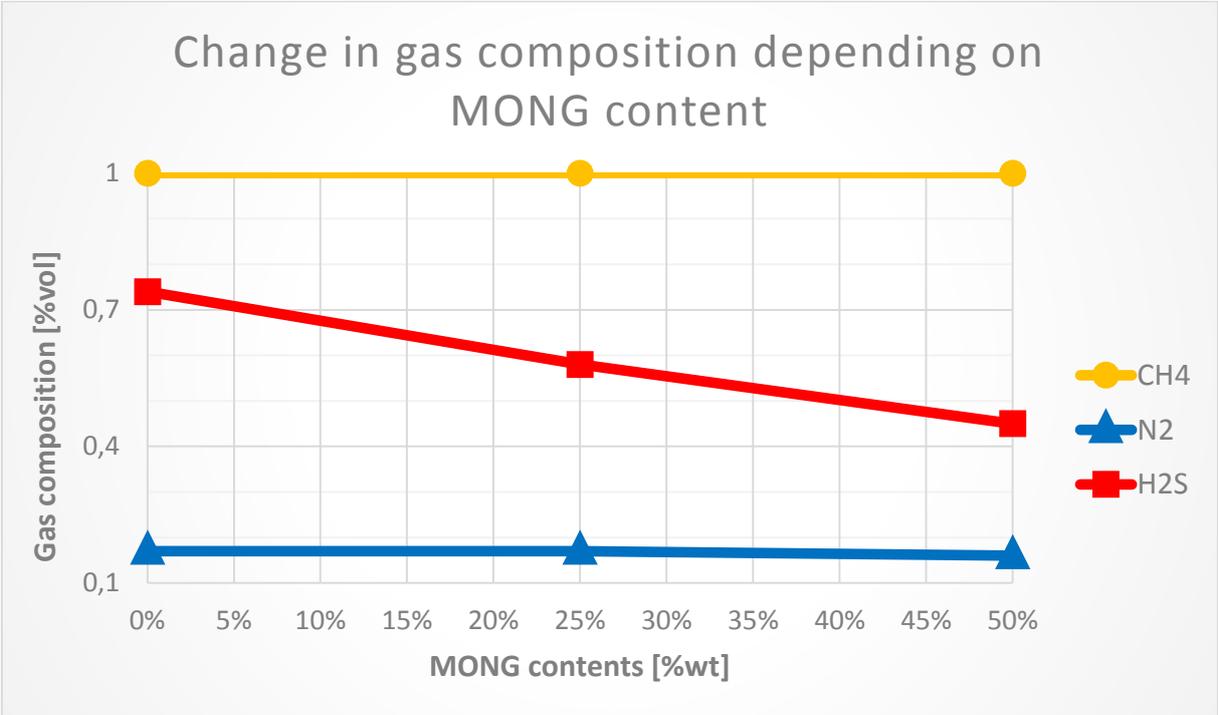


Figure 19. Illustration of change in SG composition for CH₄, N₂, and H₂S depending on CG quality and content of MONG.

An increase of MONG content shows an increase in concentration of CO and H₂ as well as a decrease in CO₂ and H₂S. This might be due to the lower amount of oxygen and increased amount of carbon and hydrogen content in the fuel thus an increase in the total HHV (see Table 19) which will have an effect on the conversion rate in the gasification process.

5.2 Techno-economic analysis

In this section the resulting cost for the investment as well as material, energy, and labour will be handled. Figures used and assumptions made in the calculations can be seen in section 4.2.

5.2.1 Material and energy flow costs and efficiency

The calculated cost for all material and energy flows using the costs described in section 4.2.1 is presented in Table 33. All prices presented are the initial value, for each year this value is increased by 1.165 % of the previous value, according to the calculated annual CPI increase in Sweden from year 2002 to 2014. Also note that material and energy costs in year one, two, and three are only at 60, 80, and 95 % due to a start-up delay to full capacity.

Table 33. Initial annual material and energy costs depending on case, where negative figures is an output of the BFP, thus a revenue. Prices are later adjusted according to calculated CPI increase of 1.165 % per year. Units in M€/year.

Case	1	2	3	4	5	6	7
Bark/logging residues	13.3	12.7	11.9	10.7	12.8	10.9	8.5
CG	-	-	-	-	6.1	23.4	54.7
Electricity	6.3	6.9	8.1	9.4	6.9	9.7	14.7
FR	-	11.4	25.6	43.9	-	-	-
Methanol	-75.2	-92.6	-114.5	-142.6	-94.9	-152.1	-258.2
O ₂	11.4	13.1	15.1	17.8	13.1	18.0	26.7
Lime kiln fuel	-3.3	-4.2	-5.3	-6.7	-4.0	-6.0	-6.6

When analysing figures for each material, a clear trend shows that the dominating post is the methanol selling revenues. However, this is highly dependent on the current selling price for methanol, which makes it vulnerable to major shifts in market price for methanol. Other posts that have quite high uncertainties are the ones for FR and CG due to assumptions of compositions and cost estimations. The price for the additional fuel feedstock will set the minimum selling cost for methanol; a high feedstock price will result in a higher methanol price and vice versa to maintain a reasonable payback of the investment.

5.2.2 Equipment and labour costs

As previously described, no individual costs for the equipment can be presented due to confidential information. Instead, the CAPEX for each case is calculated and summarized as total investment cost. The CAPEX and its corresponding depreciation value for each case are presented in Table 34. The depreciation cost for the investment is calculated using equation (3-10) and (3-11) and is further described in section 3.3. Methodology and scale parameters used for calculations are presented in section 4.2.2.

Table 34. CAPEX and depreciation costs for different cases.

Case	CAPEX [M€]	Depreciation [M€/year]
1	345	32.3
2	367	34.4
3	392	36.7
4	448	42.0
5	366	34.3
6	446	41.8
7	583	54.6

When introducing the additional fuel feedstock to the system, thus increasing fuel handling and methanol output, the higher the investment cost. The lower cost for CG compared to FR when having the same blend ratio (%wt.) is due to the lower amount of acidic components in the SG when using CG, thus a lower cost for the AGR-unit.

Labour cost, which is a cost parameter in OPEX is presented as a total sum, with and without the payroll tax in Table 35. The methodology and data for calculation of labour costs is handled in section 4.2.2. The values presented are the initial cost at the total technical lifetime of 25 years for the BFP and due to the assumption that there is an increase in salaries of 1.165 % on annual basis from CPI increase, this is added to the total OPEX. Labour costs are the same for year one, two, and three, despite not being at full capacity.

Table 35. Initial annual operating cost for labour depending on case. Prices are later adjusted according to calculated CPI increase of 1.165 % per year. Units in M€/year.

Case	1	2 to 6	7
Labour cost total	3.2	3.4	3.7
Including payroll tax	4.4	4.7	5.0

This shows that labour costs are not negligible, but a small portion compared to other costs such as oxygen and additional fuel feedstock, see Table 33.

5.2.3 Summary

When all costs have been calculated, each cost is grouped and summarized. All investment costs are placed under CAPEX, all input material and energy streams, and labour costs are grouped in OPEX, and output material streams ends up in revenues. The annual cash flow is the income that the investment will bring after depreciation and tax, PBT is the average number of years it takes for the investment to pay off, and NPV shows how much money the investment will generate under its technical lifetime if the money were brought to present time. The IRR is the profitability of the investment. Results from economic calculations are presented in Table 36, for more details see Appendix 5 – Exemplary calculation sheet for economic analysis. Depending on which type of business and/or corporation different levels of IRR are acceptable. A figure above 15 % will be considered a good investment. If NPV is above zero, the investment is considered economically feasible.

Table 36. Summary of techno-economic analysis of BFP in Vallvik, with methanol price at 560.8 €/tonne. OPEX, revenues, and cash flow are averaged figures over the technical lifetime expectancy on annual basis.

Case	CAPEX [M€]	OPEX [M€/year]	Revenues [M€/year]	Cash flow [M€/year]	PBT [years]	NPV [M€]	IRR [%]
1	345	40.5	89.5	46.2	8.3	356	11.5
2	367	55.7	110.4	51.1	7.9	408	11.9
3	392	74.7	136.5	57.2	7.5	476	12.5
4	448	98.6	170.3	66.1	7.4	553	12.7
5	366	49.9	112.8	57.6	7.0	505	13.6
6	446	76.2	180.3	91.4	5.4	936	17.6
7	583	125.1	302.0	151.1	4.3	1 700	21.9

Depending on which fuel feedstock is added, there is a difference in the IRR for each investment. The IRR of the investment increases at higher blend rates for both feedstocks, but has a much higher increase for CG. To further illustrate the potential for each fuel feedstock blend, the selling price for methanol with an IRR of 15 % for each investment is shown in Figure 20.

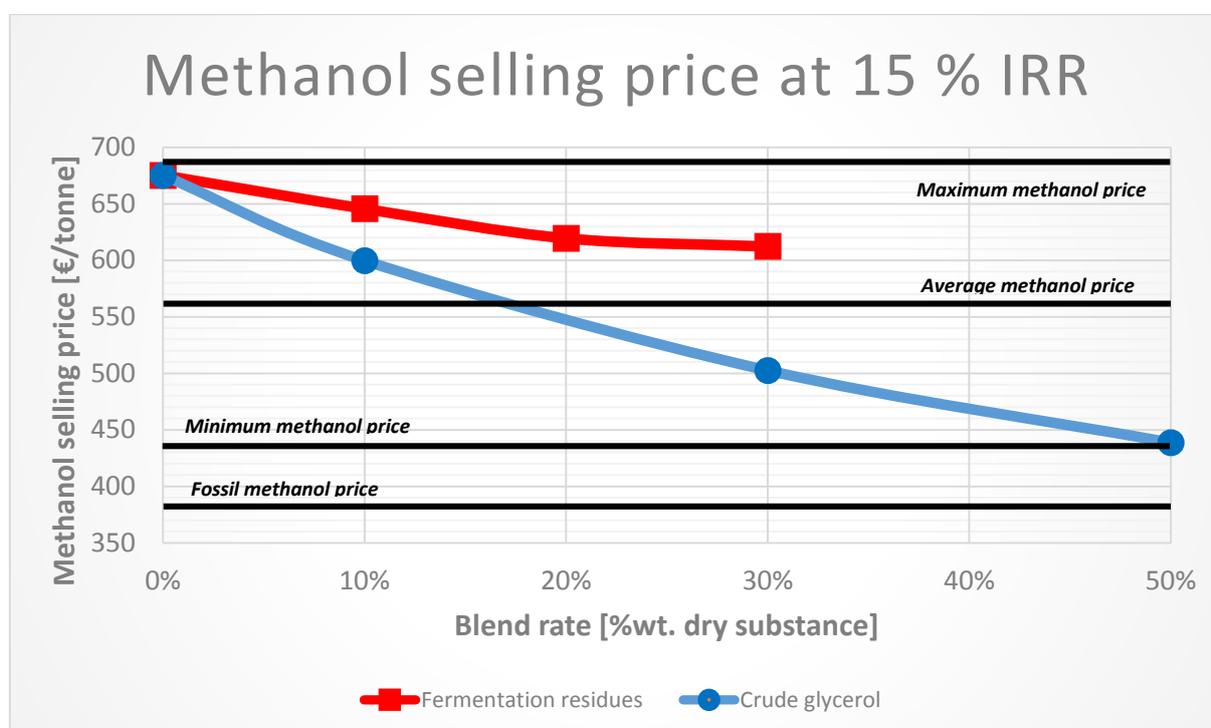


Figure 20. Minimum selling price of methanol to maintain an IRR of 15 % for different fuel blend fractions, in %wt. At 0 %wt. blend rate, only BL is used as feedstock.

The figure shows an ability to maintain an IRR of 15 % with a declining methanol selling price at increased blend rates for both CG and FR, with a steeper slope for CG. This gives CG a greater advantage to sustain a good financial statement in times when methanol prices are low. This is mainly due to the price assumptions used where FR as a fuel feedstock is more expensive than CG. To achieve an IRR of 15 % using only BL as a feedstock in the gasification will generate a minimum selling price of methanol at 675.2 €/tonne or 121.0 €/MWh. For FR at a blend rate of 30 %wt. the minimum price is at 612.2 €/tonne (109.7 €/MWh), and for CG at 50 %wt. blend rate 438.7 €/tonne (78.6 €/MWh). This shows that CG can charge less than the average methanol price, even down to the minimum price of 436.1 €/tonne and still maintain a good profitability.

A summary for different purities of CG with MONG concentrations of 0 % and 50 % is also performed. This is compared with the default concentration of MONG, shown in Table 37.

Table 37. Summary of techno-economic analysis of BFP in Vallvik, with methanol price at 560.8 €/tonne for CG blend medium with different MONG concentration. For comparative purposes, figures for default concentration of MONG in CG is shown in red. OPEX, revenues, and cash flow are averaged figures on annual basis.

Case	CAPEX [M€]	OPEX [M€/year]	Revenues [M€/year]	Cash flow [M€/year]	PBT [years]	NPV [M€]	IRR [%]
CG medium, 0 % MONG	406 (446)	67.8 (76.2)	157.5 (180.3)	79.4 (91.4)	5.6 (5.4)	794 (936)	16.8 (17.6)
CG medium, 50 % MONG	458 (446)	84.4 (76.2)	202.9 (180.3)	102.9 (91.4)	4.9 (5.4)	1 097 (936)	19.2 (17.6)

What has to be considered is that the increased purity in CG, which is lower MONG content also has a negative impact on the price. Pure grade CG can reach up to costs as high as 694.3 €/tonne (119.3 €/MWh), which is not considered in the calculations. Lower grade CG with high impurity has showed a greater conversion efficiency, lower purchase price, and an improved economic performance, but this study does not consider the amount of water that might be present in that CG. This is assumed to be removed using waste heat from various BFP processes; the cost for fuel handling, such as investment costs for drying of feedstock is included in the costs for the EFG.

The maximum acceptable cost for additional fuel feedstock, without reducing the IRR below the value attained for the reference case using BL only is evaluated and presented in Figure 21 for FR and in Figure 22 for CG.

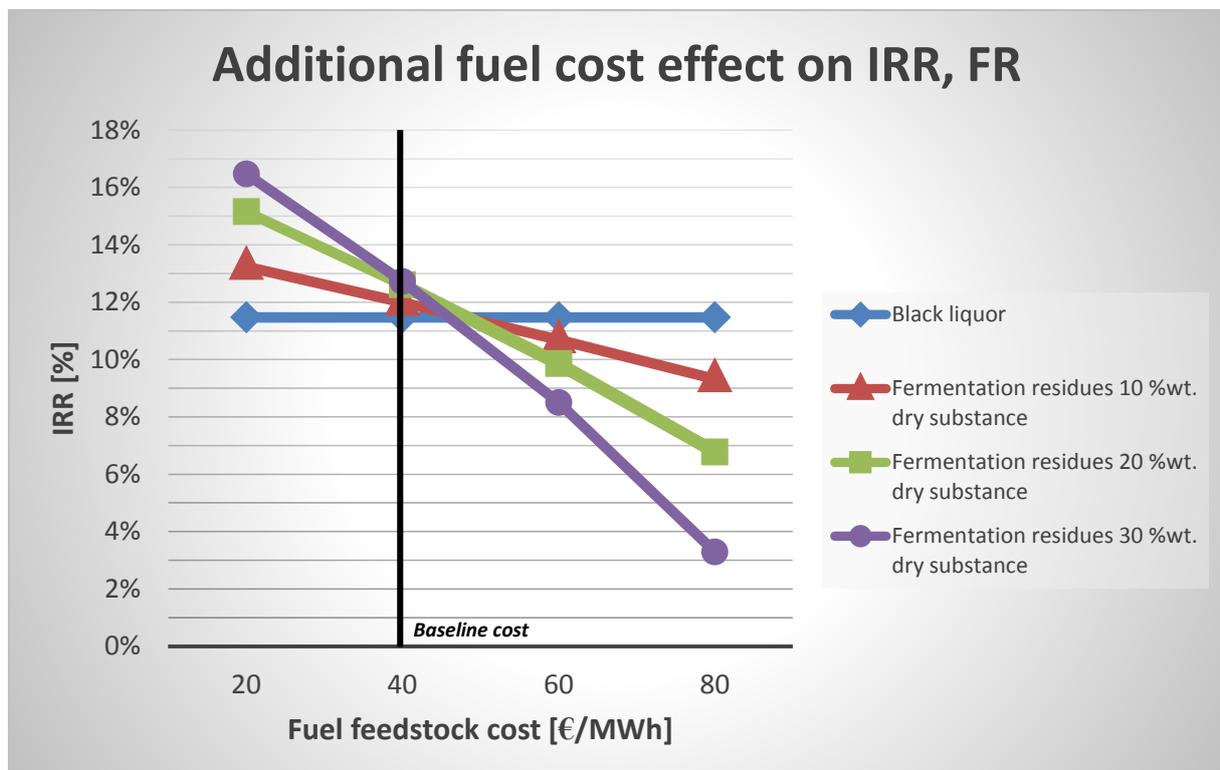


Figure 21. Effects on IRR depending on FR costs at different blend rates, with methanol price at default value.

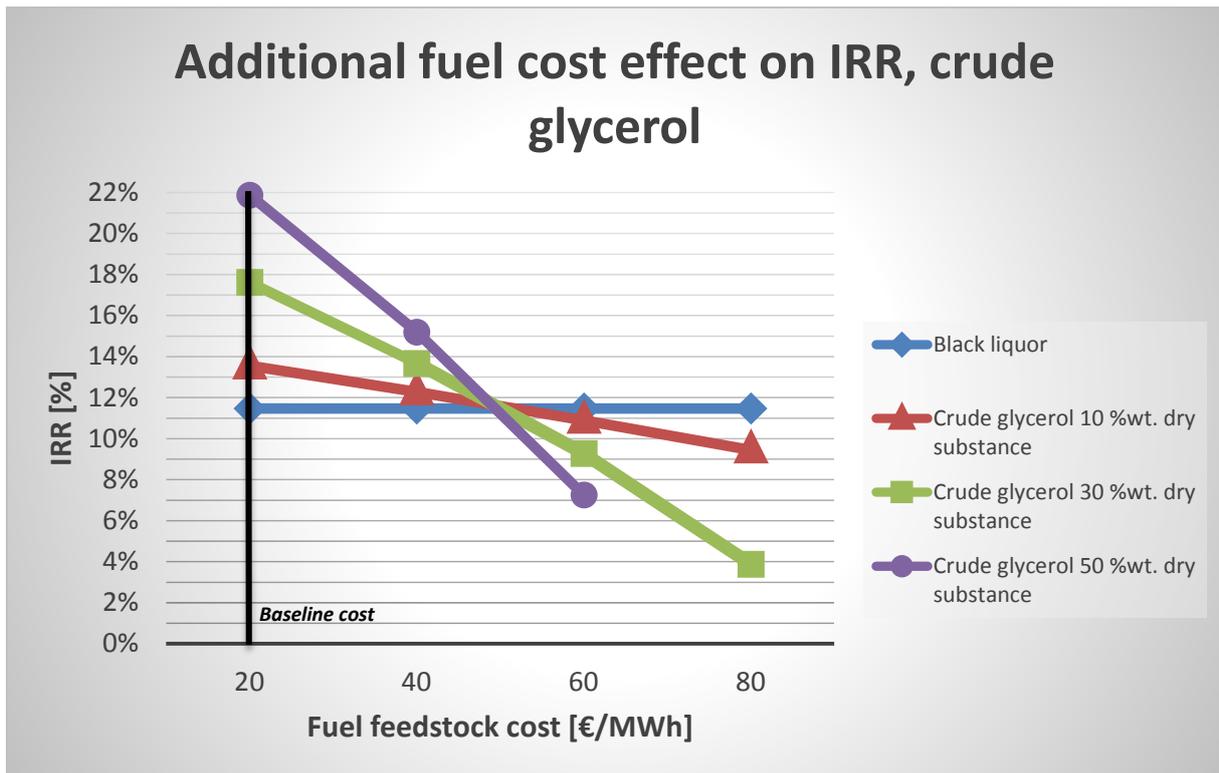


Figure 22. Effects on IRR depending on CG costs at different blend rates, with methanol price at default value.

This shows a small advantage for CG, where the maximum acceptable cost is 50.5 €/MWh for blend rate of 10 %wt. while the highest acceptable cost for FR is 46.2 €/MWh at 30 %wt. This is mainly due to the different composition of the fuel feedstock, where FR generates a SG with higher concentration of acid gas components than CG and requires a larger AGR-unit to accommodate this difference, thus a higher CAPEX and lower methanol production for FR at same blend rate as for CG. Compared to the actual baseline costs for each feedstock, CG has the greatest difference between the baseline and maximum cost. This gives CG greater financial advantage compared to FR when price for feedstock will increase.

At the given default cost for methanol the plant will break even, that is having an IRR of zero, at certain levels depending on blend, which is shown in Figure 23.

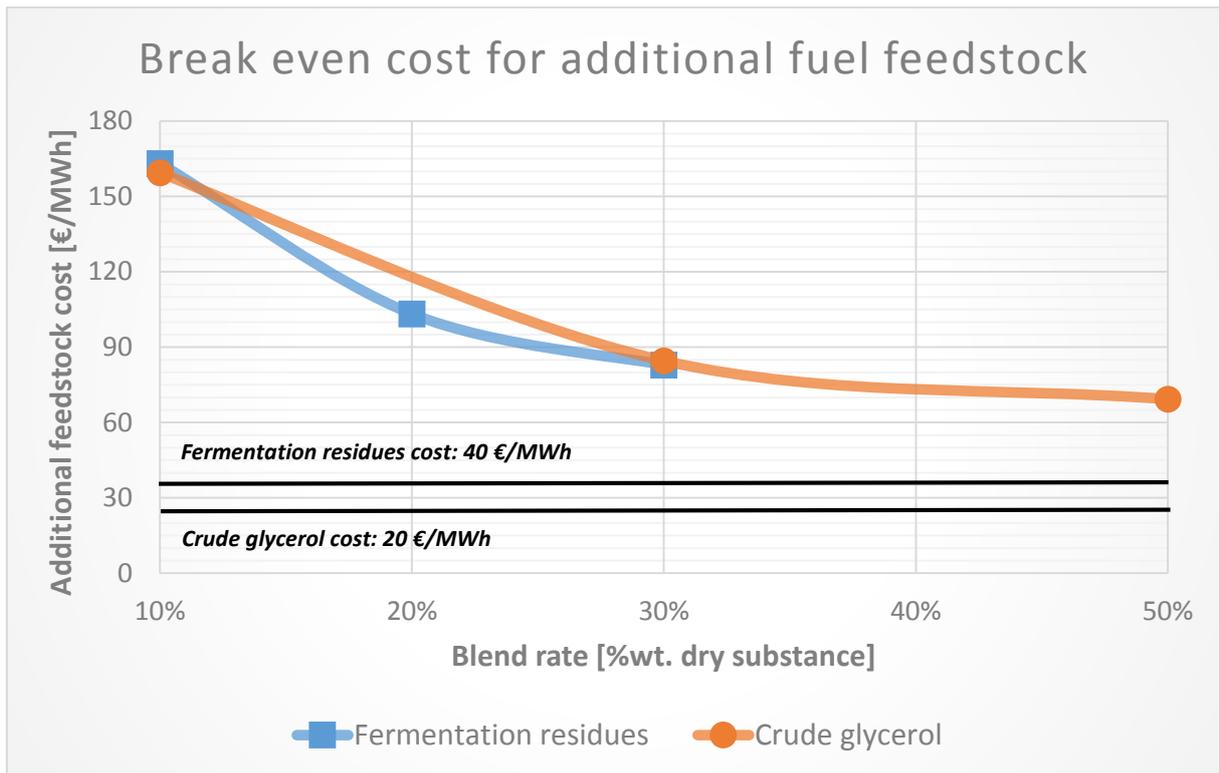


Figure 23. Maximum fuel feedstock cost at IRR of 0 % and default methanol price.

At high blend rates of CG the highest possible cost without having a negative cash flow is 69.3 €/MWh, for high blend rates of FR (30 %wt.) the maximum price is at 82.9 €/MWh. This can be compared to 84.6 €/MWh for CG at the same blend rate. The slight difference in curvature for FR and CG is possibly due to the lack of data for blend rate of 20 %wt. for CG. However, neither of the curves are close to the default cost for corresponding feedstock used in this thesis, which is why a break-even situation is considered not an issue.

Effects of change in methanol prices for case 7, which is when looking at figures in Table 36 the most profitable, shows a significant difference in IRR, graphical representation is shown in Figure 24. When matching the average fossil methanol price from Methanex Corporation (2015) of 381.0 €/tonne, the IRR of the BFP investment ends up in 11.3 %.

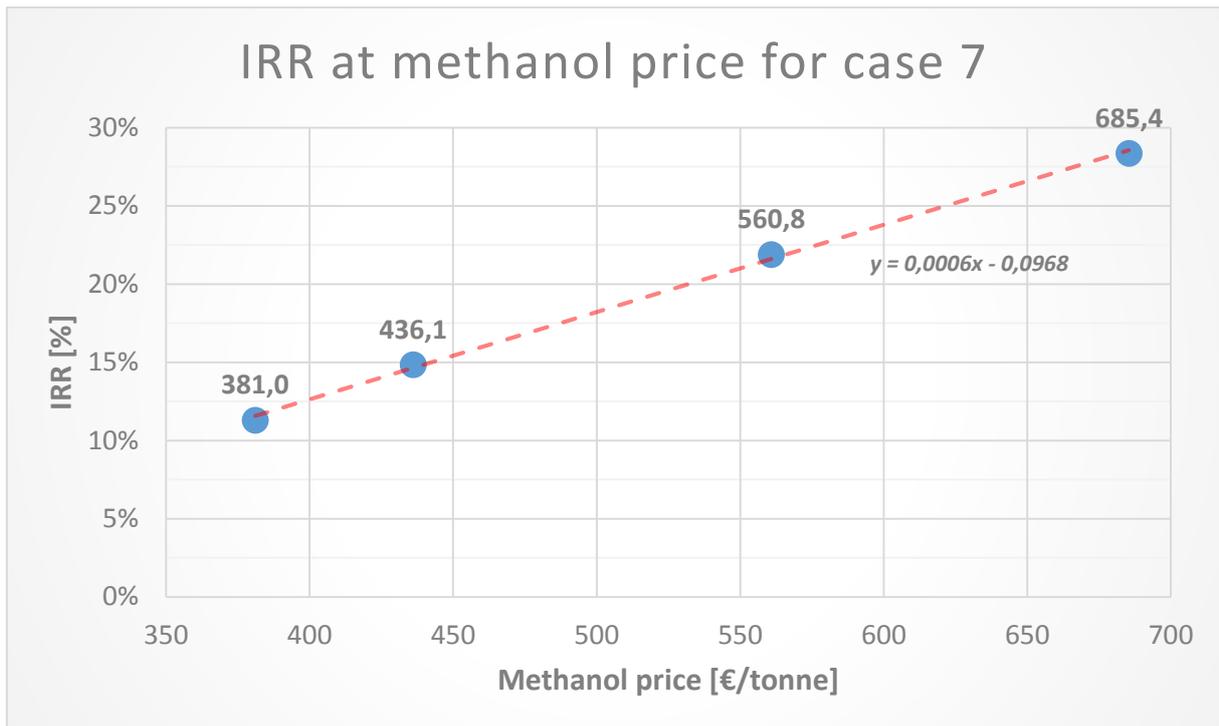


Figure 24. IRR at maximum, average, minimum, and fossil methanol prices for case 7. The dashed line represent the linear estimation of the given data, including the equation.

This demonstrates a capability to meet up with current prices on fossil derived methanol with a reasonable financial statement.

5.2.4 Sensitivity analysis

Performing a sensitivity analysis reveals how a predetermined change in any of the cost parameters will affect the outcome, in this case the IRR and NPV of case 7. This is presented in Figure 25 and Figure 26, where 0 % represents the default values previously described.

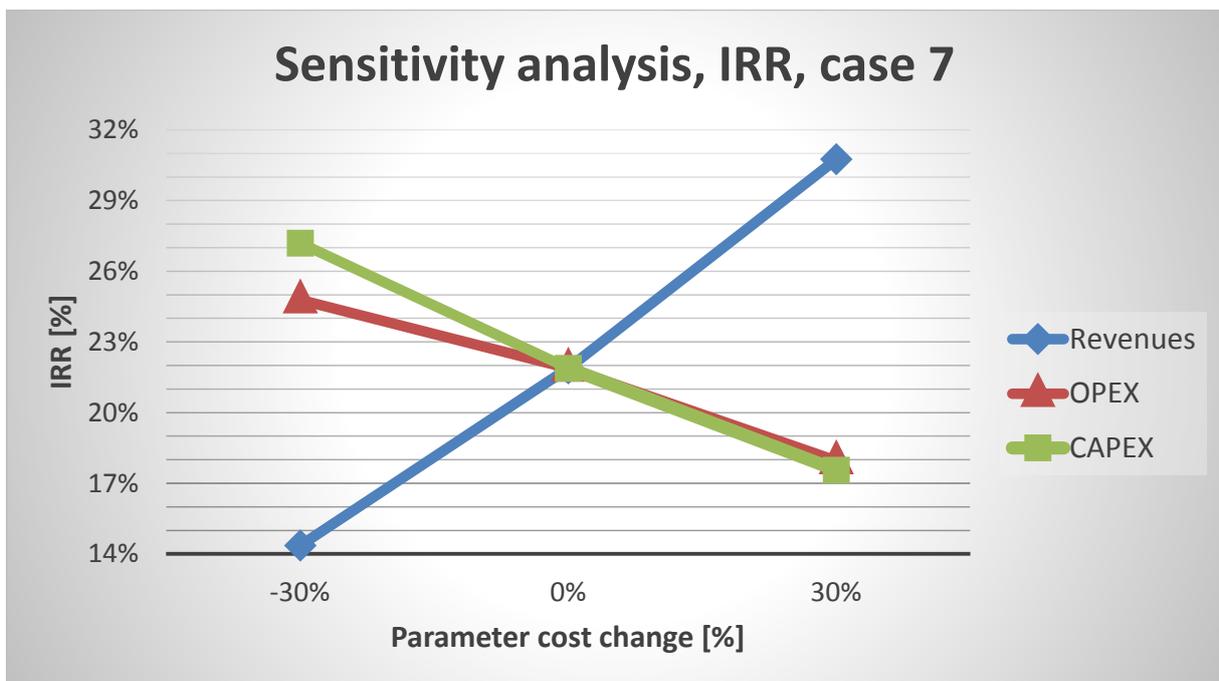


Figure 25. A sensitivity analysis of the investment IRR of case 7 with a change of $\pm 30\%$ for revenues, OPEX, and CAPEX.

The IRR of the investment shows a clear relationship between revenues and the IRR. OPEX has the least amount of influence while CAPEX has detectable change for the investment profitability.

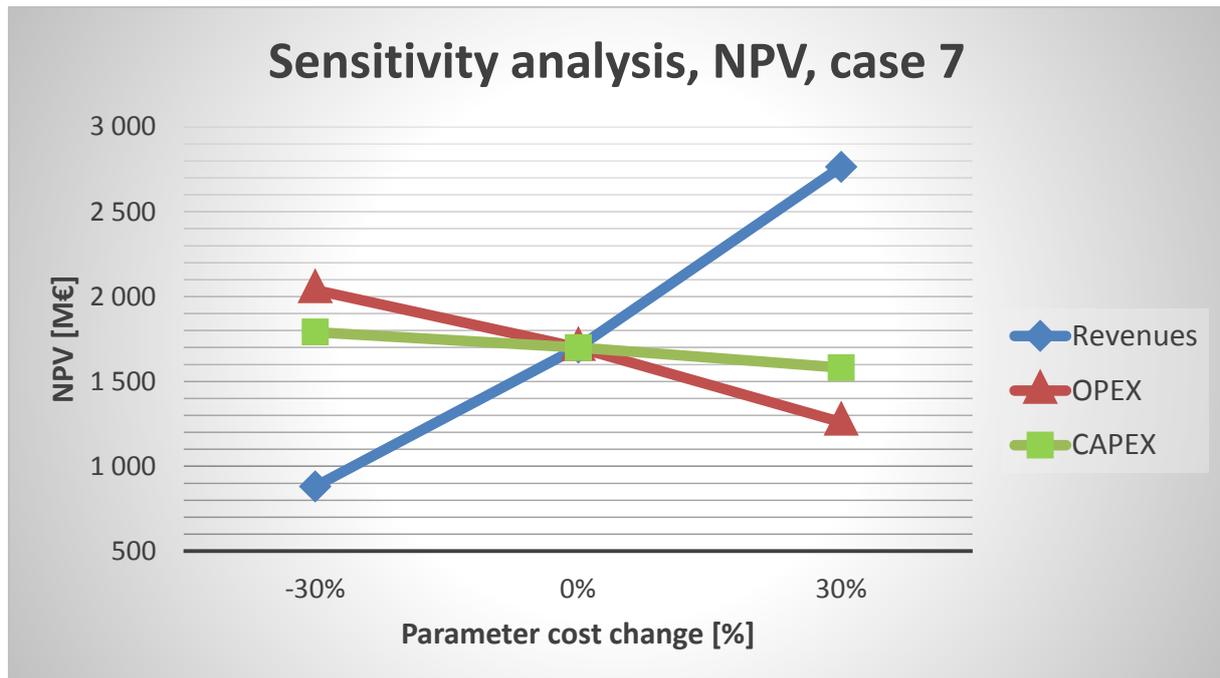


Figure 26. A sensitivity analysis of the investment NPV of case 7 with a change of $\pm 30\%$ for revenues, OPEX, and CAPEX.

Once again the group with the greatest influence, this time on NPV is revenues. However CAPEX has only a minor or barely no effect on the total outcome, due to the large amount of money generated over the investment technical lifetime expectancy compared to the initial investment cost. This will however change when looking at cheaper investments than in case 7, since NPV is dropping rapidly in these cases.

Breaking up each group into material and energy streams will reveal the post with the highest influence, thus isolating the cost parameter that needs the most attention. Sensitivity analysis performed on material/energy streams in OPEX can be seen in Figure 27 and Figure 28, and for revenues in Figure 29 and Figure 30.

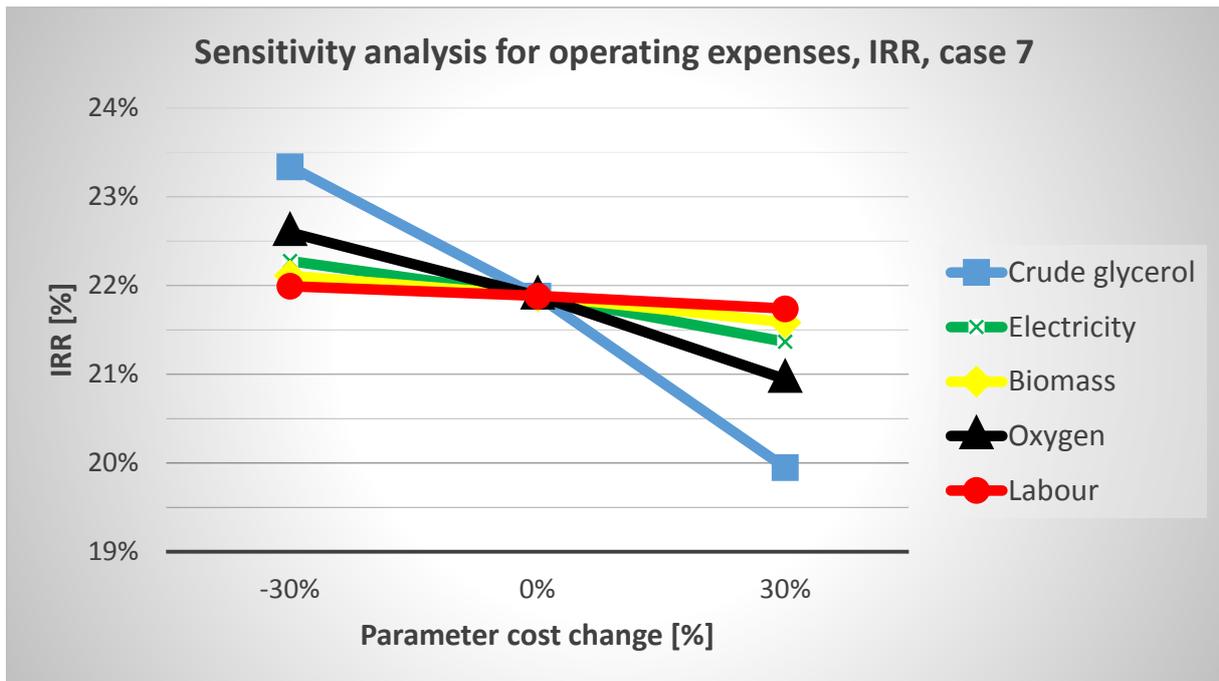


Figure 27. A sensitivity analysis of the investment IRR of case 7 with a change of $\pm 30\%$ for cost parameters in OPEX.

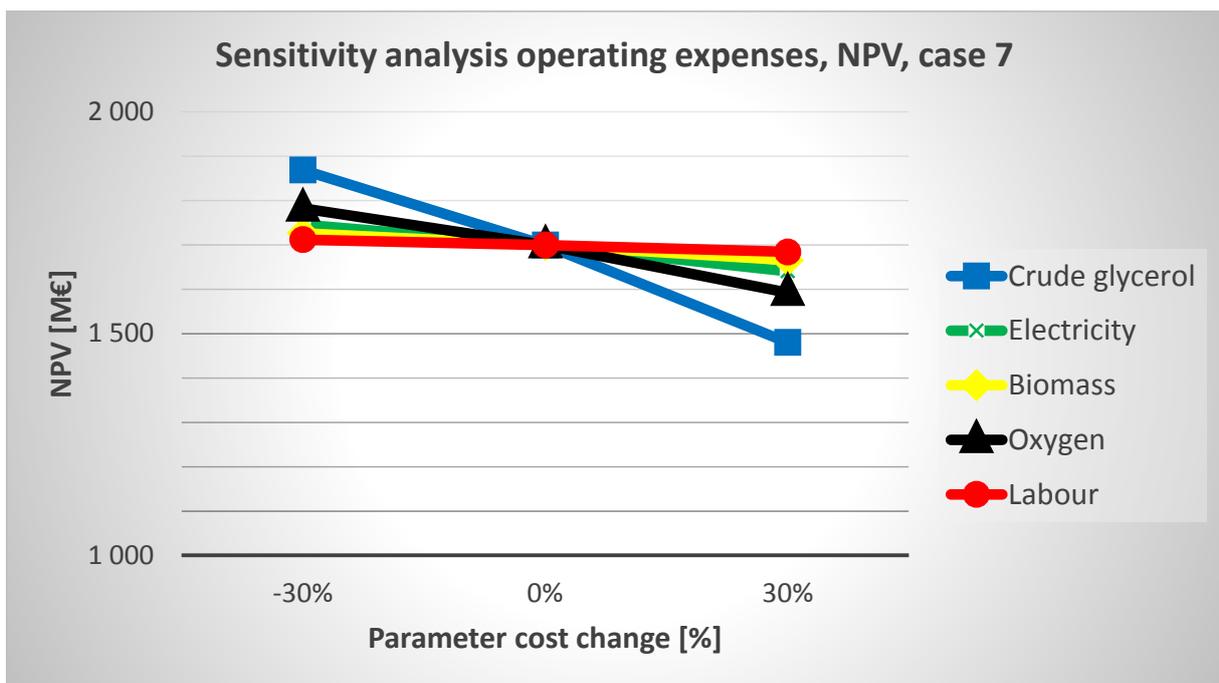


Figure 28. A sensitivity analysis of the investment NPV of case 7 with a change of $\pm 30\%$ for cost parameters in OPEX.

This shows that the cost parameter with highest effect on the OPEX is the feedstock CG for case 7, shortly followed by oxygen and electricity in that order. This is true for case 7 where 50 %wt. of the fuel feedstock comes from CG, any other case where blend is lower will affect the analysis. Lower amount of additional fuel feedstock will lower the influence of the additional feedstock as well as lowering the influence of oxygen and electricity, due to lower fuel flows and a higher steam and power production in the biomass boiler. This will instead shift towards a higher influence of biomass price on IRR and NPV. Labour has only minor influence in all cases. Same principle will also apply for using FR as feedstock, but will in general make the effect on IRR and NPV greater when modifying feedstock cost, due to a higher initial price for feedstock.

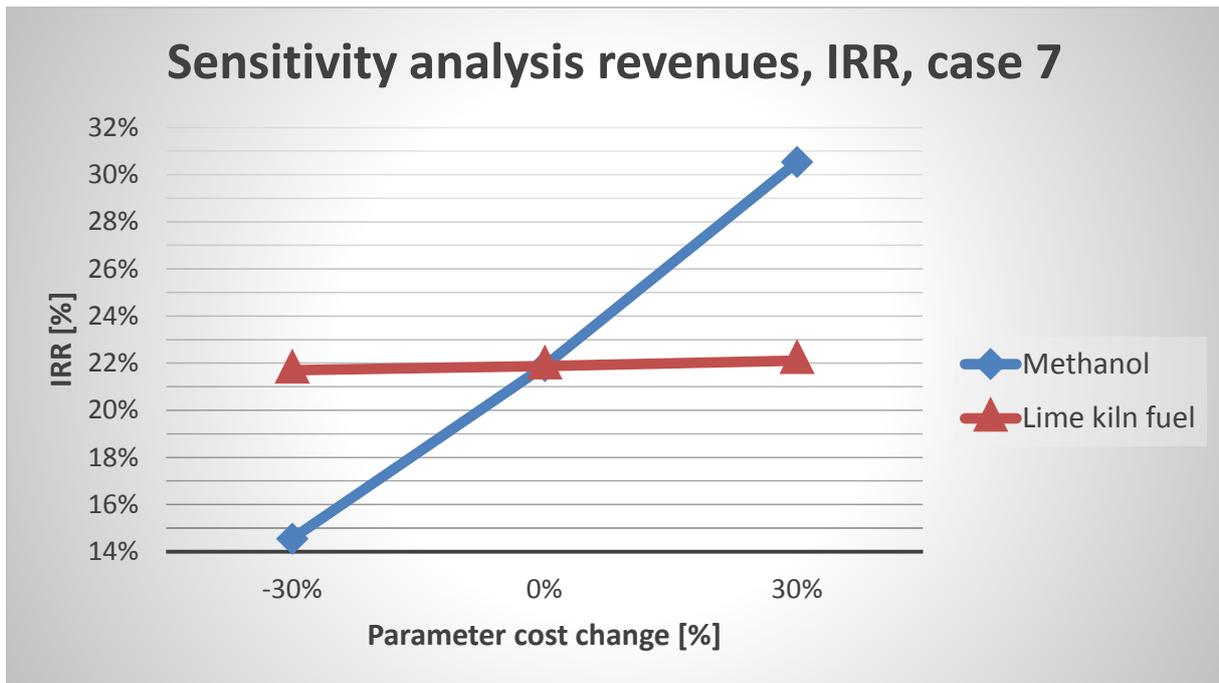


Figure 29. A sensitivity analysis of the investment IRR of case 7 with a change of $\pm 30\%$ for posts in revenues.

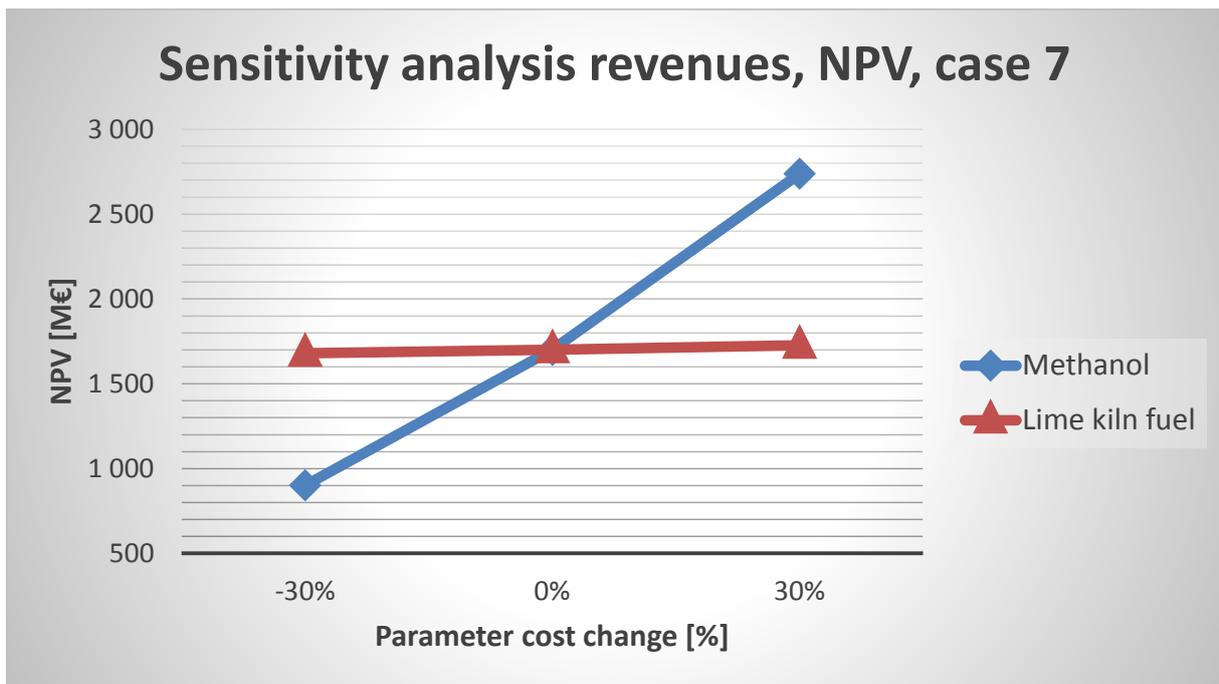


Figure 30. A sensitivity analysis of the investment NPV of case 7 with a change of $\pm 30\%$ for posts in revenues.

Here it is clear that the post with the most influence on revenues, and for the entire BFP as a whole is the pricing of methanol. The reduced cost for lime kiln fuel has barely no influence on the total outcome. This shows that it is crucial to determine the price ranges for methanol, where lower price limits are depending on the desired production capacity.

6 Conclusion

With previously described conditions, an investment utilizing BLG with additional fuel feedstock is considered profitable. Estimated NPV for an integrated BFP range from 356 to 1 700 M€ and has an IRR of 11.5 to 21.9 %, depending on the size of the investment chosen. In previous case studies performed on Vallvik PM (Andersson et al., 2013) (Stare, et al., 2012), the reference case of using BL as fuel feedstock ended up with an IRR of 17.5 %. This difference is related to the pricing for methanol (984 €/tonne compared to 560.8 €/tonne), which has been determined to be the key issue for the BFP profitability.

With a CAPEX of 583 M€, the best investment is an integrated BFP using BL and 50 %wt. of CG, which will increase methanol production by over 240 %, have an IRR of 21.9 %, a NPV of 1 700 M€, and a straight PBT of 4.3 years. Also having another commodity on the market will help lowering the risk of having a poor financial statement for the PM.

The selling price for methanol is of major importance, since it is the single most contributing revenue of the BFP. The lower limit for the selling price of methanol varies between each case and at what price the additional fuel is purchased. When compared to fossil derived methanol, a price match is still possible with a quite reasonable IRR of 11.3 % using BL with added CG of 50 %wt.

To further increase the methanol production and improving the methanol conversion rate, CG is the most suitable additive. From assumed composition and simulations, adding CG gives a greater increase in the conversion efficiency than FR, has a higher heating value, is cheaper, and generates higher IRR and NPV. An increase of the world's biodiesel production will yield a higher availability of CG, thus a greater probability of lower CG cost and an improvement of the economic feasibility of the BFP.

The blend rate of the CG should be chosen so that suppliers can provide the amount necessary for gasification, but as high as possible since it generates a greater profitability, and at a recommendable cost around 20 €/MWh. Costs temporarily above this level can be acceptable, but in a longer period of time this might impede the total profitability of the plant. Maximum cost acceptable for CG at low blend rate is 50.5 €/MWh to attain same IRR as for just using BL as gasification fuel. Due to high pricing of FR, an increased methanol production using FR as additional feedstock generated a minor increase in IRR. FR might still be an alternative if the cost for the feedstock is somewhat lower than the 40 €/MWh used for calculations in this thesis. A high pricing for both fuel feedstock can be countered by lowering production, which will help to maintain the methanol production as well as keeping a fairly reasonable IRR of the investment. The paper from Andersson et al (2013), where pyrolysis oil were used as an additional feedstock in BLG, gave an acceptable cost of 70 €/MWh for the feedstock. This is somewhat higher than the calculated figures in this thesis of 50.5 €/MWh for CG and 46.2 €/MWh for FR. The main reason to this is because of the difference in the methanol price used, 984 €/tonne compared to 560.8 €/tonne.

From simulation results it is shown that a high concentration of glycerol or high grade quality of the CG is not of importance. This is however not confirmed through pilot scale testing, which is why a conclusion on how CG quality affects plant performance cannot be done. Low grade CG has in general a lower purchase cost, which will improve the economic performance of the BFP. Do observe that this thesis has not taken into account any recovered methanol contained in the CG, doing this should further improve the profitability.

Since there is a larger amount of SG handled in the cases involving additional fuel feedstock, there is also an increase in the electric power input for the compressors for MSY process. This in combination with a lower power production due to lower HP steam production in the biomass boiler gives a modest declining energy efficiency of the methanol production with increasing weight ratios of additional fuel. However, this is compensated by the increase in conversion efficiency from fuel feedstock input to methanol output, which makes BLG with additional fuel a more profitable choice when the prices for methanol and additional fuel are within the limits used in this thesis.

Uncertainties which highly affect the results are the pricing of methanol, fuel feedstocks, and the elemental composition of the fuel feedstock mixture. From sensitivity analysis the post with the highest effect on the IRR for the BFP is the price for methanol, and with a price set based on ethanol converted on energy density there is a certain risk this can affect the economic performance of an investment. The effect of feedstock price depends on the quantity of additional feedstock, but at high blend rates (30-50 %wt.) this is a dominating OPEX post. It is also determined that fuel feedstock composition will affect the methanol output, thus a shift in a direction which lowers the methanol output will affect the BFP profitability. Fluctuating costs for fuel feedstocks and methanol, in combination with SG compositions based on results from simulation with assumptions and simplifications, is of paramount importance. Any of these should be further investigated to ensure a high reliability on the BFP profitability.

7 Future work

During the project, several issues and questions were raised which were not handled either due to lack of time or were not within the project scope, but still too interesting to be left behind without being noticed. These final thoughts and reflections on the thesis are handled in this section.

7.1 Issues

An initial problem started with the composition of CG and FR. Since there was no specific sample to be evaluated and many different studies showed all possible types of compositions, the idea of what a CG was had to be defined. Through discussion with Gustav Haggström, the definition of a CG was set which is described in section 4.1.1. The FR had the predetermined condition as a wood based source, but since no data for pine was available elemental composition and heating value of hybrid poplar was instead used.

Early on it was established by Gustav Haggström that there might be some problem with mixing of BL and FR, due to dissolving issues. This problem did develop under room temperature conditions, which might be avoided when the fuel mixture is heated. Thereby the assumptions were made that as long as the blend rate did not exceed 30 %wt. this would not create any problem, such as too large particles in the EFG. For CG there was no detectable problems with mixture of the two feedstock, which is why calculations with a higher blend rate were performed.

The cost for CG and FR was difficult to acquire, since both were fluctuating and varied depending on which study were analysed. CG also had great variety in cost due to different qualities and compositions, which is why an average cost from different studies were used. The cost for FR was even more difficult to acquire, since it currently is not a large commodity and thereby were not many studies found that could be used as a source.

A goal was previously set to investigate how absence or presence of subsidies were to affect the outcome of the BFP economically. This was however abandoned due to problem to establish which type of subsidy was to be available as well as its effect.

7.2 Improvements

To improve the work performed in this thesis, a couple of things could be further investigated. First of all an actual sample of a CG and FR should be analysed to determine an elemental composition that is closer to a real situation, instead of using based on assumptions of previous studies. A deeper understanding of the effects of the CG and FR compositions, both in technical and economical perspective is of interest.

The availability of the additional fuel feedstock should be further looked into, such as if there is a risk of shortage in supply, any limitations in transport, and similar types of issues. Thoughts like these have been raised during this thesis even though it was not in the scope of the project, but is crucial for determine the possibility to even apply a chosen case.

Since methanol recovery in any additional feedstock was not accounted for in this thesis, it would be interesting to further investigate how this methanol output would affect the final outcome. Here it is essential to determine the methanol contents in the feedstock, which varies in each sample.

A later validation of the model through pilot scale verification of co-gasification technology would be of interest. How well does all assumptions and calculations fare on the end result compared to a full scale case and what can be done to further improve the model.

The economic analysis showed a steady improvement of IRR for CG at increasing blend rates. It would be of interest to investigate how higher blend rates would affect the profitability of the plant and where the upper limit would be.

Finally the abandoned track where subsidies were included or excluded would be interesting to further investigate, if there would be a predetermined type of subsidy to use.

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9 Appendix

9.1 Appendix 1 – Aspen® modelling settings

Parameter	Setting
GC	
<i>SG inlet temperature</i>	Case: 1) 215.2°C, 2) 213.8°C, 3) 212.5°C, 4) 211.3°C 5) 213.6°C, 6) 210.4°C, 7) 207.5°C, MONG0%) 212.1°C, MONG50%) 208.8°C
<i>SG inlet pressure</i>	31 bar
<i>SG inlet mass flow</i>	Case: 1) 146 119 kg/h, 2) 167 059 kg/h, 3) 193 392 kg/h, 4) 227 434 kg/h, 5) 164 299 kg/h, 6) 216 442 kg/h, 7) 311 453 kg/h MONG0%) 212 269 kg/h, MONG50%) 220 712 kg/h
<i>SG composition</i>	See Table 38
<i>SG H₂O removal</i>	Type: Flash2 Flash type: Temperature 140°C, Pressure 0 bar
<i>Split for heat from Flash2</i>	Design specification: Sufficient to heat FW to SG temperature
<i>Heater FW</i>	Inlet stream: Condensation from SG Flash type: Pressure 0 bar
<i>Heater LPS</i>	Inlet stream: Temperature 117°C, Pressure 4.5 bar, H ₂ O Flash type: Pressure 0 bar Design specification: Outlet temperature of 150°C by modifying mass flow
<i>SG HEX2</i>	Type: HeatX Exchanger specification: Hot stream outlet temperature 40°C, Streams: SG vapour, condensation liquid. Cold stream inlet: Temperature 10°C, Pressure 1 bar, H ₂ O; Cold stream outlet: Design specification: Temperature 95°C by modifying mass flow of cold stream
WGS	
<i>Split of SG to WGS</i>	Split fraction: 0.5
<i>IPS stream</i>	Temperature: 255°C, Pressure: 29 bar, H ₂ O. Design specification: SG final composition $\frac{H_2 - 0.03 \cdot (CO + H_2 + CH_4 + N_2)}{CO + 0.03 \cdot (CO + H_2 + CH_4 + N_2)} = 2.11$ by modifying IPS mass flow
<i>MUW to pump</i>	Inlet conditions: Temperature 40°C, Pressure 1 bar, H ₂ O Outlet conditions: Pressure 41 bar Design specifications: Has equal mass flow to water in waste streams from WGS, AGR, and PDU
<i>Heater after mixer</i>	Temperature 200°C, Pressure -1 bar
<i>HEX after heater</i>	Type: MHeatX Hot stream: Pressure 0 bar Cold stream: Specification: Temperature 265°C, Pressure 0 bar
<i>Reactor</i>	Type: REquil Flash type: Pressure 0 bar, Duty 0 cal/sec Reactions: See equation (1-1)
<i>H₂O separator</i>	Type: Flash2 Flash type: Temperature 35°C, Pressure 22.2 bar
<i>Waste water split</i>	Split fraction: 0.01
<i>IPS steam supply</i>	Design specification: Mass flow of steam should match the mass flow required from the WGS model, diverted from split of IPS steam produced in the MSY process.

Rectisol™ “black box”	
<i>Type</i>	Sep2
<i>H₂S concentration in clean SG</i>	Split fraction: Design specification: H ₂ S is 0.1 ppm of mole fraction in clean SG
<i>CO₂ concentration in clean SG</i>	Split fraction: Design specification: $\frac{CO_2}{CO_2+H_2+CO} = 0.04$
<i>Outlet flash</i>	Clean gas: Pressure -3.96 bar, Temperature change -11°C H ₂ S and CO ₂ stream: -1.67 bar, Temperature change 3°C
<i>Energy to AGR</i>	Design specification: MPS, LPS and work is diverted from splits so that they satisfy the working condition parameters for the Rectisol™.
Methanol synthesis	
<i>SG compressor</i>	Discharge pressure: 106 bar
<i>HEX prior to MSY</i>	Type: HeatX Exchanger specification: Cold stream outlet temperature 150°C Cold stream inlet: Temperature 117°C, Pressure 4.5 bar, H ₂ O Hot stream outlet temperature: Design specification: 166°C by modifying cold stream mass flow
<i>Reactor</i>	Type: REquil Flash type: Temperature 261°C, Pressure -8 bar Reactions: See equation (1-1), (1-2) and (1-3)
<i>Pump for IPS</i>	Inlet stream: Temperature 117°C, Pressure 14 bar, H ₂ O Discharge pressure: 29 bar
<i>Heater for IPS</i>	Type: Heater Flash type: Pressure 0 bar Design specification: Outlet stream temperature 255°C by modifying mass flow into pump
<i>Split for IPS</i>	Split fraction: Design specification: Mass flow of IPS for WGS
<i>Valve for IPS</i>	Calculation type: Adiabatic flash for specified outlet pressure (pressure changer), Outlet pressure 14 bar
<i>HEX after MSY</i>	Type: MHeatX Hot stream inlet: Pressure 0 bar Cold stream inlet: Temperature 117°C, Pressure 4.5 bar, H ₂ O Cold stream outlet: Specification: Temperature 150°C, Pressure 0 bar Hot stream outlet temperature: Design specification: 130°C by modifying cold stream mass flow
<i>Cooler after HEX</i>	Type: Heater Flash type: Temperature 38°C, Pressure -3 bar
<i>Condenser</i>	Type: Flash2 Flash type: Pressure -12 bar, Duty 0 cal/sec
<i>Tail gas split</i>	Split fraction: 0.04 (case 7: 0.02)
<i>Recycled methanol compressor</i>	Discharge pressure: 106 bar
<i>Separator, crude methanol</i>	Type: Flash2 Flash type: Pressure 3.44505 bar, Duty 0 cal/sec

PDU	
<i>Distiller</i>	Type: RadFrac Calculation type: Equilibrium Number of stages: 15 Condenser: Total Reboiler: Kettle Valid phases: Vapour-liquid Convergence: Standard Distillate rate: 300 kmol/hour Reflux ratio: 1.5 Feed streams: Stage 12, Above-Stage Product streams: Methanol, Stage 1, Liquid; Condensate, Stage 15, Liquid Pressure: Top/Bottom, Top stage/Condenser pressure 3.03975 bar Design specification: Mass recovery, Target: 0.99, Components: Methanol, Product stream Vary: Adjusted variable: Distillate rate, lower bounds 10 kmol/hour, upper bound 7 000 kmol/hour
<i>Energy to reboiler</i>	Design specification: LPS steam mass flow is diverted from split so the energy from 150°C of LPS steam down to 115°C of feedwater is sufficient to provide for reboiler energy.
AGE	
<i>Steam to AGE</i>	LPS steam is constant at 1 000 kg/hour and is diverted from split
Pumps	
<i>Efficiency</i>	0.8
<i>Drivetrain</i>	0.99
Compressors	
<i>Isentropic efficiency</i>	0.82
<i>Mechanical efficiency</i>	0.95
Turbines	
<i>Isentropic efficiency</i>	0.87
<i>Mechanical efficiency</i>	0.98
HPS from boiler	
<i>Stream</i>	Temperature 470°C, Pressure 61 bar, H ₂ O Design specification: Mass flow modified to satisfy remaining MPS and LPS steam demand for BFP and PM.

Table 38. Gas composition from SIMGAS simulations after EFG, before GC.

Gas component	CG medium 50% MONG	CG medium 0% MONG	FR high	FR medium	FR low	CG high	CG medium	CG low	BL only
CO	0.155	0.126	0.135	0.126	0.116	0.161	0.141	0.118	0.106
H2	0.176	0.150	0.145	0.139	0.132	0.186	0.163	0.138	0.124
CH4	0.004	0.004	0.004	0.004	0.003	0.004	0.004	0.003	0.003
CO2	0.061	0.076	0.081	0.081	0.081	0.061	0.069	0.077	0.080
N2	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001
H2S	0.002	0.003	0.003	0.003	0.003	0.002	0.002	0.003	0.004
H2O	0.601	0.641	0.631	0.647	0.664	0.586	0.620	0.660	0.682

9.2 Appendix 2 – Feedstock composition calculation sheet

Flow	
Mass flow black liquor [tonne/day DS]	1500
Mass flow additional feedstock [%wt. DS]	50,00%
Total mass flow [tonne/day DS]	3000
Mass flow additional feedstock [ton/dag TS]	1500
LHV BL [MJ/kg DS]	12,13
HHV [MJ/kg DS]	13,56
Total mole flow [kmol/day]	25355,4141
Crude glycerol	
HHV [MJ/kg DS]*	16
*Wei et al. 2011	
HHV [MJ/kg DS]**	22,89
**From "Heat of formation"	

Biomass to EFG	
Mass flow [kg/s]	34,72222222
Effect HHV [MW]	632,7243991
SG after GC	
Mass flow SG post GC [kg/s]	36,61219444
mH2 [%mass]	4,83658%
nH2 [kmol/s]	0,878416667
nCO [kmol/s]	0,764413889
nCH4 [kmol/s]	0,019544111
CO2 [MJ/kmol]	-393,52
H2O [MJ/kmol]	-241,83
CO [MJ/kmol]	-110,53
CH4 [MJ/kmol]	-74,87
SG after GC	
Effect LHV [MW]	444,4294247
LHV [MJ/kg]	12,13883602
HHV [MJ/kg]	13,21981165
Effect HHV [MW]	484,0063147
Methanol	
Mass flow [kg/s]	17,01483333
HHV MeOH [MJ/kg]	22,884
Effect HHV [MW]	389,367446
LHV MeOH [MJ/kg]	20,094
Effect LHV [MW]	341,896061

Conversion efficiency	
Cold gas efficiency	80,25824%
Feedstock to methanol	61,74203%
SF LHV [MJ/kg DS]	15,94088623
Effect DS [MW]	553,5029942

Composition black liquor	%wt, Dry]	M [g/mol]	[g/kg]	n [mol/kg]
C	32,50%	12,0107	325	27,05920554
H	3,40%	1,00794	34	33,7321666
S	5,52%	32,066	55,2	1,72144951
O	35,96%	15,9994	359,6	22,47584284
Na	20,23%	22,98977	202,3	8,799566068
K	2,18%	39,0983	21,8	0,557568999
Cl	0,13%	35,453	1,3	0,036668265
N	0,08%	14,00674	0,8	0,05711536
Summa	100,0%			
DS [%]	73,2%			
HHV [MJ/kg DS]	13,56			
Molar mass [kg/kmol]	94,43958319			
Mole flow [kmol/dag TS]	15883,17048			
Mole fraction [%]	0,626421261			

Composition additional fuel	%wt, Dry]	M [g/mol]	[g/kg]	n [mol/kg]
C	47,1%	12,0107	470,97185	39,21268961
H	9,3%	1,00794	92,982065	92,2496033
S	0,0%	32,066	0	0
O	41,7%	15,9994	417,16755	26,07394934
Na	1,9%	22,98977	18,878539	0,821171267
K	0,0%	39,0983	0	0
Cl	0,0%	35,453	0	0
N	0,0%	14,00674	0	0
Summa	100,0%			
DS [%]	94,5%			
HHV [MJ/kg DS]	22,89			
Molar mass [kg/kmol]	158,3574135			
Mole flow [kmol/day DS]	9472,24362			
Mole fraction [%]	0,373578739			

Composition total	%wt, Dry]	M [g/mol]	[kg]	n [mol]
C	39,80%	12,0107	1193957,8	99407842,73
H	6,35%	1,00794	190473,1	188972654,8
S	2,76%	32,066	82800	2582174,266
O	38,84%	15,9994	1165151,3	72824688,28
Na	11,06%	22,98977	331767,81	14431106
K	1,09%	39,0983	32700	836353,4987
Cl	0,07%	35,453	1950	55002,39754
N	0,04%	14,00674	1200	85673,04026
Sum	100,0%			
DS [%]	82,50%			
HHV [MJ/kg DS]	18,22246269	LHV [MJ/kg DS]	16,365871	
Molar mass [kg/kmol]	118,3179256			
Mole flow [kmol/day DS]	25355,4141			
Mole fraction [%]	1			

9.3 Appendix 3 – Boiler calculations to determine biomass flow

Calculation of stoichiometric combustion were performed using parameters shown in Table 39 and Table 40. Results of this were used as input for the boiler calculations, which is shown in Table 41. Fuel mass flow output changes according to steam mass flow input.

Table 39. Settings for air supplied to the stoichiometric combustion of biomass.

Parameter	Value
Lambda	1.2
Air temperature	30°C
Relative humidity	40 %

Table 40. Component weight distribution of biomass fuel used in biomass boiler.

Component	Value [%wt.]
C	28.4
H	3.3
O	22.7
N	0.2
S	0.0
Ash	6.1
H ₂ O	39.3

Table 41. Calculation sheet of mass flow of fuel to biomass boiler, based on steam production. Data in red indicates a constant value of a parameter, all else is calculated based on stoichiometric combustion calculations and conservation of mass and energy for the boiler.

Parameter	Value	Parameter	Value
Tref [°C]	0	<u>Steam</u>	
		Tcond [°C]	115
<u>Air</u>		Pcond [bar]	61
l [kmol/kg,fuel]	0.1433	Tv [°C]	470
Cpa [kJ/kg,K]	29.09	Pv [bar]	61
Ta [°C]	30	Δh [kJ/kg]	2 863
ha [kJ/kmol]	872.7	m [kg/s]	19
<u>Flue gases</u>		Q [MW]	53.3
g [kmol/kg,fuel]	0.1804		
Cpg [kJ/kg,K]	31.79		
Tg [°C]	130		
hg [kJ/kmol]	4132.3		
<u>Ash</u>			
mr/mf	13.00%		
Cpr [kJ/kg,K]	0.9		
Tr [°C]	300		
hr [kJ/kg]	270		
<u>Efficiencies</u>			
ηcomb	99.00%		
Ploss	1.50%		
<u>Fuel (GROT)</u>			
Hfuel [MJ/kg]	8.5		
Tfuel [°C]	30	14	MJ/kg,DS
Cpfuel [kJ/kg,K]	2.3	39.30%	moisture content
hfuel [kJ/kg]	69		
mf [kg/s]	6.92		
Cfuel [€/MWh]	18.56		
Power [MW]	59		
Fuel to steam efficiency	90.7%		

9.5 Appendix 5 – Exemplary calculation sheet for economic analysis

Year	CAPEX	OPEX	Revenue	Depreciation	Net income	Tax	Net income	Cash flow	NPV	IRR
0	-582 953 466 €							-582 953 466 €	-582 953 466 €	-582 953 466 €
1		-68 530 938 €	160 754 312 €	-54 610 369 €	37 613 005 €	-7 177 722 €	30 435 283 €	85 045 652 €	81 774 666 €	69 776 858 €
2		-90 738 079 €	216 836 795 €	-54 610 369 €	71 488 347 €	-14 630 297 €	56 858 050 €	111 468 419 €	103 058 819 €	75 036 130 €
3		-108 038 973 €	260 494 290 €	-54 610 369 €	97 844 948 €	-20 428 749 €	77 416 199 €	132 026 568 €	117 371 138 €	72 918 750 €
4		-114 775 561 €	277 399 844 €	-54 610 369 €	108 013 914 €	-22 665 922 €	85 347 993 €	139 958 362 €	119 636 994 €	63 421 429 €
5		-116 113 050 €	280 632 408 €	-54 610 369 €	109 908 989 €	-23 082 838 €	86 826 151 €	141 436 520 €	116 250 510 €	52 584 534 €
6		-117 466 125 €	283 902 642 €	-54 610 369 €	111 826 147 €	-23 504 613 €	88 321 534 €	142 931 903 €	112 961 159 €	43 599 844 €
7		-118 834 968 €	287 210 983 €	-54 610 369 €	113 765 646 €	-23 931 303 €	89 834 344 €	144 444 712 €	109 766 110 €	36 150 699 €
8		-120 219 762 €	290 557 878 €	-54 610 369 €	115 727 746 €	-24 362 965 €	91 364 782 €	145 975 151 €	106 662 613 €	29 974 592 €
9		-121 620 693 €	293 943 773 €	-54 610 369 €	117 712 711 €	-24 799 657 €	92 913 054 €	147 523 423 €	103 648 000 €	24 853 904 €
10		-123 037 950 €	297 369 125 €	-54 610 369 €	119 720 807 €	-25 241 438 €	94 479 369 €	149 089 738 €	100 719 685 €	20 608 230 €
11		-124 471 721 €	300 834 393 €	-54 610 369 €	121 752 303 €	-25 688 367 €	96 063 936 €	150 674 305 €	97 875 155 €	17 088 007 €
12		-125 922 201 €	304 340 042 €	-54 610 369 €	123 807 472 €	-26 140 504 €	97 666 968 €	152 277 337 €	95 111 975 €	14 169 248 €
13		-127 389 583 €	307 886 543 €	-54 610 369 €	125 886 590 €	-26 597 911 €	99 288 680 €	153 899 049 €	92 427 781 €	11 749 158 €
14		-128 874 065 €	311 474 371 €	-54 610 369 €	127 989 937 €	-27 060 647 €	100 929 290 €	155 539 659 €	89 820 278 €	9 742 519 €
15		-130 375 846 €	315 104 008 €	-54 610 369 €	130 117 794 €	-27 528 775 €	102 589 019 €	157 199 388 €	87 287 240 €	8 078 677 €
16		-131 895 126 €	318 775 942 €	-54 610 369 €	132 270 447 €	-28 002 359 €	104 268 088 €	158 878 457 €	84 826 507 €	6 699 058 €
17		-133 432 112 €	322 490 666 €	-54 610 369 €	134 448 185 €	-28 481 461 €	105 966 724 €	160 577 093 €	82 435 983 €	5 555 096 €
18		-134 987 007 €	326 248 677 €	-54 610 369 €	136 651 301 €	-28 966 147 €	107 685 154 €	162 295 523 €	80 113 634 €	4 606 529 €
19		-136 560 023 €	330 050 481 €	-54 610 369 €	138 880 089 €	-29 456 480 €	109 423 609 €	164 033 978 €	77 857 485 €	3 819 974 €
20		-138 151 368 €	333 896 587 €	-54 610 369 €	141 134 850 €	-29 952 528 €	111 182 323 €	165 792 692 €	75 665 620 €	3 167 753 €
21		-139 761 258 €	337 787 513 €	-54 610 369 €	143 415 886 €	-30 454 356 €	112 961 530 €	167 571 899 €	73 536 180 €	2 626 917 €
22		-141 389 908 €	341 723 780 €	-54 610 369 €	145 723 503 €	-30 962 031 €	114 761 472 €	169 371 841 €	71 467 360 €	2 178 440 €
23		-143 037 537 €	345 705 916 €	-54 610 369 €	148 058 011 €	-31 475 623 €	116 582 388 €	171 192 757 €	69 457 409 €	1 806 546 €
24		-144 704 365 €	349 734 457 €	-54 610 369 €	150 419 723 €	-31 995 200 €	118 424 523 €	173 034 892 €	67 504 627 €	1 498 155 €
25		-146 390 618 €	353 809 942 €	-54 610 369 €	152 808 956 €	-32 520 831 €	120 288 125 €	174 898 494 €	65 607 364 €	1 242 420 €
Average		-125 068 753 €	301 958 615 €	-54 610 369 €	122 279 492 €	-25 804 349 €	96 475 143 €	151 085 512 €		
								NPV	1 699 890 826 €	0 €
CPI increase annually	1.165%							IRR	21.9%	
Tax rate	22%							PBT	3.9	
Real interest rate	4%									
Year	Production capacity									
1	60%									
2	80%									
3	95%									
4	100%									