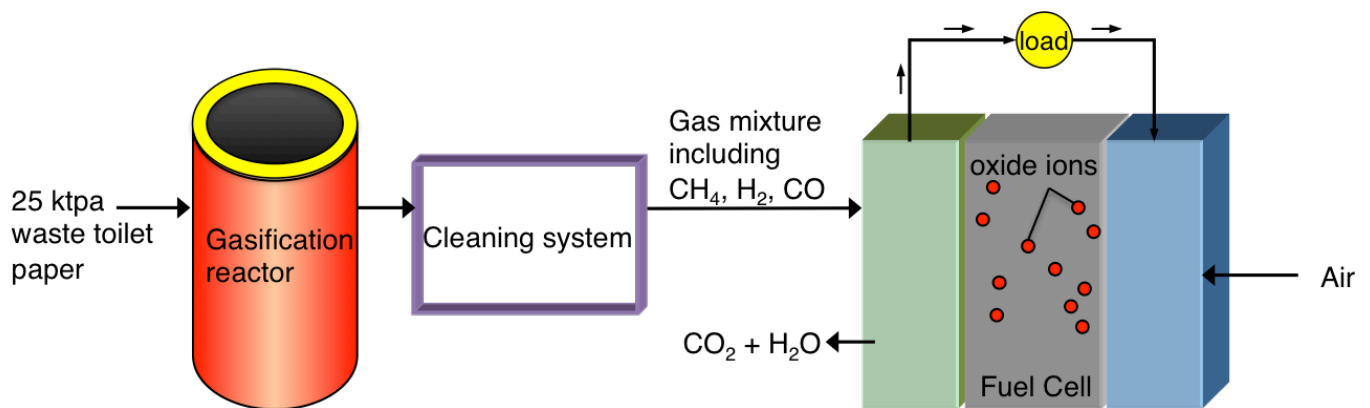# Waste Toilet paper to Electricity –
A techno-economic feasibility study

Master Thesis report

Name: Els van der Roest
Student number: 4220021
Email: elsvdroest@gmail.com
Daily supervisor: ir. Mijndert van der Spek (UU)
Internal supervisor: dr. ir. Andrea Ramirez (UU)
External supervisor: prof. dr. Gadi Rothenberg (UvA)
Date: 11-03-2016

Universiteit Utrecht

Abstract

The goal of this research was to assess the techno-economic feasibility of a waste toilet paper (WTP) to electricity plant. This concept uses new technologies to create local electricity out of a waste source. The specific case was to convert WTP from the wastewater treatment plant from Waternet in Amsterdam-West to electricity at the site of the waste incineration company AEB that lies directly next to it. Input would be 25.000 tonne WTP per year with 60% moisture and an energy content of about 5MWth, delivered at the gate of the AEB. The system set up would include a gasification reactor and a high temperature fuel cell.

The method for assessing the system had six steps. First, the elemental composition and energy content of the fuel were determined. The second step was to design a system for conversion of WTP to electricity. A gasification model was created in excel to determine the product gas composition. Third, the electricity yield and energy efficiency of the system were calculated. Fourth, an economic analysis was performed to calculate the Net Present Value (NPV), Internal Rate of Return (IRR), discounted Pay Back Period (PBP) and Levelized Cost of Electricity (LCOE). In the fifth step learning effects on the gasifier and fuel cell system were assessed. Finally, an uncertainty analysis was conducted which included both a sensitivity analysis and a pedigree analysis.

The results section starts with the elemental composition of WTP followed by the product gas composition after gasification. It continues with a system design that includes (among other components) the indirect MILENA gasifier from ECN and a solid oxide fuel cell (SOFC). Further, the energy analysis reveals an electricity production of 2.8 MW at 57.2% electrical efficiency and total efficiency of 69.7%. Next, the economic results show that the system is (currently) not economically feasible; NPV is -38.2 Meuro, IRR -24.2% and LCOE 23.05 eurocent/kWh. The payback period could not be calculated. Subsidy could reduce the LCOE to 12.67 eurocent/kWh, while learning effects of the SOFC system could reduce the LCOE to 15 eurocent/kWh.

Finally, the uncertainty analysis showed that priorities for future research are the SOFC investment cost and the water content and amount of WTP. Although the system is currently unfeasible, it shows that a waste source can be converted to electricity at high efficiency. The main recommendations are to invest in SOFC development and scale up the project.

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Glossary

AEB   Afval Energie Bedrijf
AFC   alkaline fuel cell
CC   carbon conversion
CEPCI   Chemical Engineering Plant Cost Index
CHP   combined heat and power
CPI   Consumer Price Index
DMFC   direct methanol fuel cell
ECN   Energie Onderzoekscentrum Nederland
FC   fuel cell
HDS   hydrodeshulphurization
HHV   higher heating value
IRR   internal rate of return
LCOE   levelized cost of energy/electricity
LHV   lower heating value
MCFC   molten carbonate fuel cell
NPV   net present value
PAFC   phosphoric acid fuel cell
PBP   payback period
PEMFC   proton exchange membrane fuel cell
SDE+   stimulering duurzame energieproductie
SNG   synthetic natural gas
SOFC   solid oxide fuel cell
STOWA   Stichting Toegepast Onderzoek Waterbeheer
WTP   waste toilet paper
WWTP   waste water treatment plant

Executive summary

Introduction and research goal

Toilet paper is a product that is essential for most people living in the Western world. An average European person consumes around 10-14 kg each year (Ruiken, Klaversma, Breuer, & Neef, 2010). Altogether we provide waste (water) treatment companies with an almost continuous supply of toilet paper. In the Netherlands, toilet paper is flushed and arrives at a waste water treatment plant (WWTP) where it is processed and later on becomes part of the waste-stream.

The water company Waternet is currently looking for new ways to process the waste toilet paper (WTP), as it could improve the energy balance of their installations. Besides, WTP still contains 70-80% cellulose, which is valuable material. Waternet is investigation several options, such as fermenting of the WTP to biogas or fatty acid production. In this study a route is proposed to convert the WTP into electricity by means of a gasification process followed by electricity production in a high temperature fuel cell. In this way, the (continuous) waste supply is matched to a continuous demand, thereby also supporting local electricity production and increasing sustainability and security of energy supply.

Case and research goal

The case investigated in this research is an input stream of 25.000 tonnes WTP/year (60% moisture) from the WWTP at Amsterdam-West, which can be used directly as feed for an installation on the site of the waste incineration company (AEB). The installation contains a gasifier, gas cleaning system and a high temperature fuel cell (Figure 1).
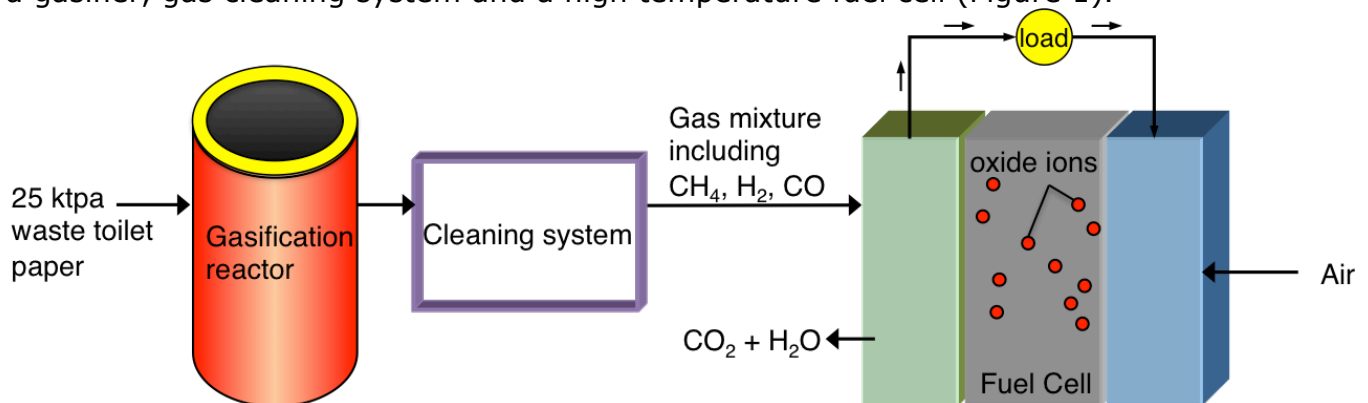
Figure 1 Illustration of the three-step process for converting waste toilet paper to clean electricity. Ktpa= kilo tonne per annum.

Preliminary calculations showed that around 6400 households can be provided with electricity from WTP, and the scale of the installation will be around 5MWth. Both the AEB and Waternet are interested in the possibility of building a WTP to electricity plant, but need more information about the feasibility of the project before they can make an investment decision, which leads to the research goal:

The goal of this study is assessing the techno-economic feasibility of a waste toilet paper to electricity plant.

To achieve the goal, the following research questions were formulated:
1. *What could be a possible process design for a WTP to electricity plant?*
2. *How would such a process perform in terms of energy efficiency and electricity yield?*
3. *What is the economic feasibility of the process?*
4. *What are the largest sources of uncertainty in the results?*

Method

Before the actual method was designed, first the system boundaries were defined. The case considered is a WTP to electricity plant, built on the AEB site with a feed-stream of the WWTP at Amsterdam-West of 25.000 tonne/year wet WTP (60% moisture). The system boundaries are crossed at the point where the WTP arrives with 60% moisture at the AEB, and at the point where the electricity that is produced leaves the system. Besides, there can be some heat-exchange with the streams on the AEB site when necessary. This means that the extraction of WTP from waste water is outside the system boundaries. When the system boundaries were set, six steps were performed that together form the method.

1) <u>Fuel characterization</u>: Because WTP is not a regular fuel, experiments and calculations were performed to determine the elemental composition and energy content.
2) <u>System design</u>: Starting from the basic system components, which are the gasifier, cleaning equipment and fuel cell, a more detailed design was created. First, a suitable gasifier was chosen and modelled in excel to find out the producer gas composition. Then an investigation of the fuel cell characteristics and inlet requirements was performed. Based on the difference between the gasifier output and required fuel cell input, the necessary cleaning equipment was chosen. During the process of system design, interviews with experts from ECN and a gasifier and cleaning system supplier (Royal Dahlman) were conducted.
3) <u>System analysis</u>: For every system component, mass and energy balances were constructed in excel. Subsequently, a (simplified) heat integration was done to match heat sources and sinks. Then, a total energy balance was created and the energy efficiency and electricity yield of the process were calculated.
4) <u>Economic analysis</u>: Data for the economic analysis were gathered via different sources. The gasifier and cleaning system prices could be obtained directly from a supplier. The fuel cell costs were obtained from literature. The investment cost were combined with other factors to calculate the Total Plant Cost. The cost data were input for the calculation of four economic indicators: Net Present Value (NPV), Internal Rate of Return (IRR), discounted Pay Back Period (PBP) and levelized cost of electricity (LCOE). The NPV gives the present value of a project when all current and future costs-flows are taken into account. The IRR is the rate of return one can expect within the project lifetime. The PBP indicates from which moment on the generated benefits of the project can be seen as profits. Finally the LCOE gives the costs of electricity production over the project lifetime. Together they give a good indication of the economic feasibility of the system. Lastly, the possibility of subsidy was investigated based on the SDE+ subsidy scheme.
5) <u>Learning effects</u>: Because both biomass-gasifiers and fuel cells are not (fully) commercialized yet, effects of learning were examined. Learning describes the effect of decreasing costs when the production expands. Learning rates and cumulative installed capacity were derived from the literature.
6) <u>Uncertainty</u>: To check how sensitive the LCOE is for changes in input parameters, a sensitivity analysis was done. A selection of thirteen input parameters and sensitivity ranges was made. Besides, a pedigree analysis was performed in which the same parameters were scored by a group of researchers from Utrecht University plus experts from companies. Each parameter received a score on four criteria (proxy, reliability of source, completeness and validation process) on a scale of 0-4. The results of the sensitivity analysis and the pedigree analysis were combined in a diagnostic diagram to indicate which parameters need most attention in further research.

Results and discussion
Fuel characterization
Compared to wood, the WTP contains more ash (4.75%) and relatively more O, due to the high cellulose content: 70-80% in WTP versus 15-30% in wood (Nagel, 2008). The energy content used in further calculations is 16.13 MJ/kg (Lower Heating Value) and 17.49 MJ/kg (Higher Heating Value).

System design
The designing process started by selecting an appropriate gasifier based on the scale (5MWth) of the system, the fuel flexibility and low tar and nitrogen content of the gas. The gasifier of choice is the indirect MILENA gasifier, developed by ECN (van der Meijden, 2010), as it fits all the requirements. The fuel cell of choice is a Solid Oxide Fuel Cell (SOFC), which is chosen because it can withstand some contaminants, can use hydrogen, carbon monoxide and methane as fuel and is used more often in combination with a gasifier system.

The total system design is shown in Figure 2 and starts with a dryer. Then after the gasifier there is a cyclone, flue gas filter and OLGA tar removal system that recycles tars to the gasifier where they are burned. After the OLGA system a water scrubber removes hydrogen chloride (HCl) and part of the ammonia (NH_3) from the gas and reduces the water content. Then a compressor increases the gas pressure to 3 bars for treatment in the hydrodesulphurization (HDS) reactor, where all organic sulphur is converted to hydrogen sulphide (H_2S). Besides, the HDS converts HCN to NH_3 and methane (CH_4) and hydrocarbons with double bonds are hydrogenated. Then a zinc oxide (ZnO) reactor is added to remove the H_2S to < 1 ppm levels. Finally the gas is depressurized and fed to the fuel cell. On many places in the system, heat exchangers are installed to heat or cool the gas or inlet stream to the required temperature. With this design, the WTP with 60% moisture can be gasified, cleaned to SOFC inlet requirements and converted to electricity.
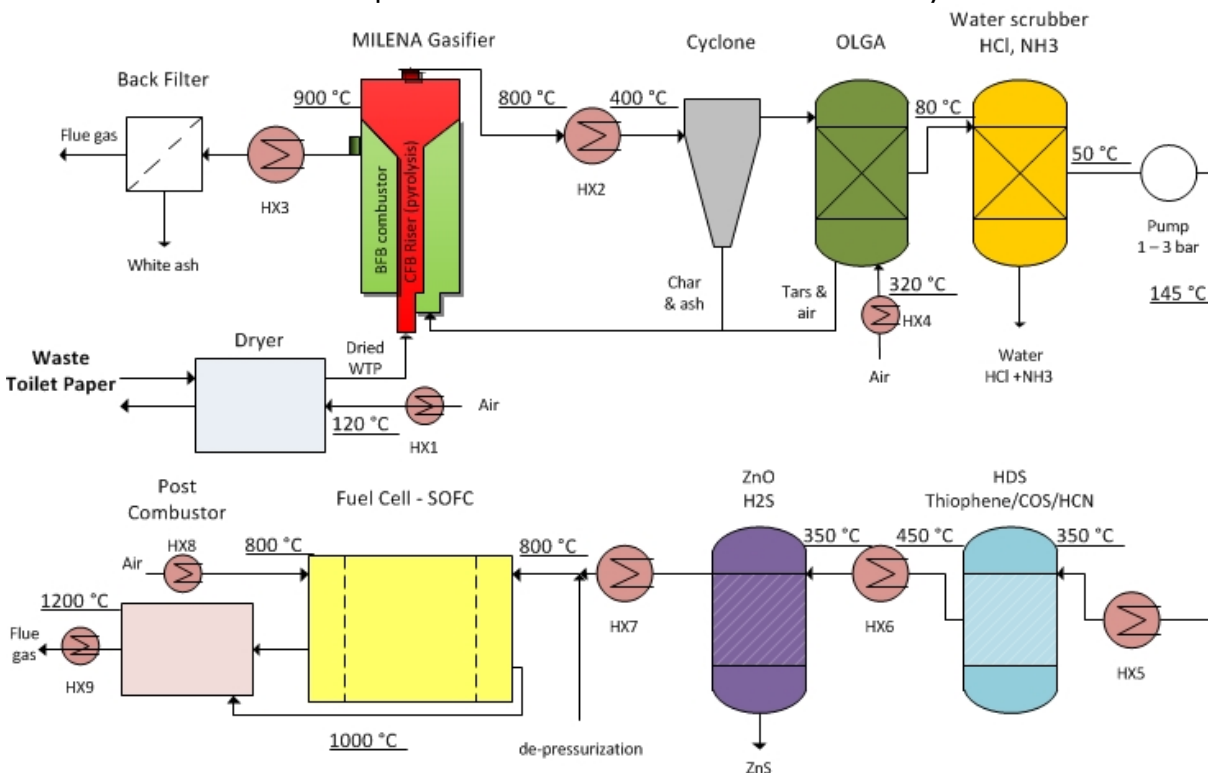
Figure 2 Final system design

Energy analysis
Overall the system will produce 2.8 MW electricity and 0.55 MW heat. The energy efficiency of the total process is 69.7% and the electricity yield is 57.2%. This is a high efficiency when compared to other options (see Graph 1) such as waste incineration with a maximum of 30% electrical efficiency. It performs better than coal/biomass integrated gasification combined cycle systems (39-45% electrical efficiency) and is comparable to natural gas combined cycle

systems (56% elec. eff.). Other investigated biomass gasification systems with an SOFC showed (slightly) lower efficiency of 40-54%, which could be due to the scale of the system, energy integration options and the age of the studies (around 2009).
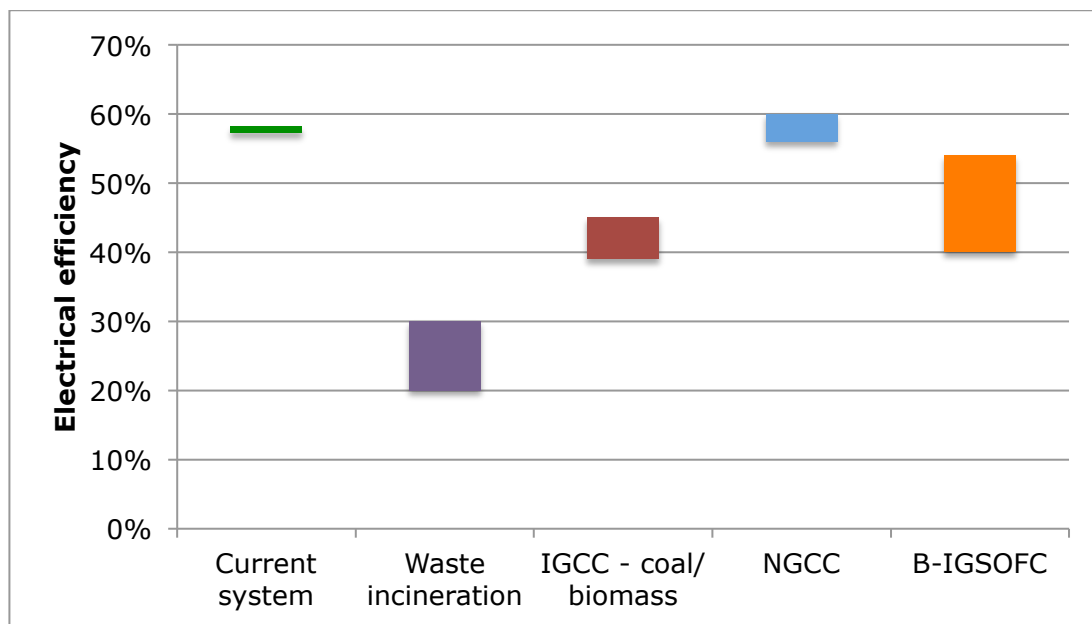
Graph 1 Comparison of WTP to electricity with other electricity production technologies. Green is the current system for WTP to electricity as discussed in this report. Purple is waste incineration, with an electrical efficiency varying between 20-30% (European Commission, 2006; Gemeente Amsterdam & AEB, 2012). Red is Integrated Gasification combined cycle for a combination of coal and biomass (van den Broek, Hoefnagels, Rubin, Turkenburg, & Faaij, 2009), Blue is Natural Gas Combined Cycle (van den Broek et al., 2009; Graus, Roglieri, Jaworski, & Alberio, 2008; Thattai, Wittebrood, Woudstra, Geerlings, & Aravind, 2014) and orange shows the efficiency range of different scales of Integrated biomass gasification SOFC – Gas turbine (B-IGSOFC) – some wit gas turbine- systems (Aravind, Woudstra, Woudstra, & Spliethoff, 2009; Jin, Larson, & Celik, 2009; Nagel, Schildhauer, & Biollaz, 2009).

Economic analysis and learning
The NPV for the system is -38.20 million euro, with an IRR of -24.2% and an LCOE of 23.05 eurocent/kWh, the payback period could not be calculated. These results make the system economically unfeasible. With subsidy of 10.38 eurocent/kWh the system can still not reach break-even with an NPV of -22.03 Meuro, an IRR of -9.7%, and an LCOE of 12.67 eurocents/kWh.
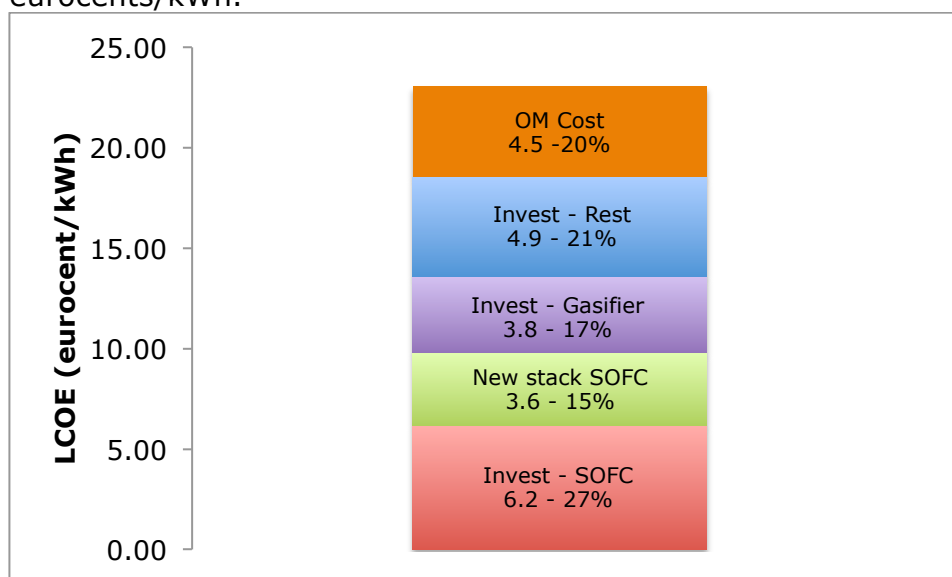
Graph 2 Build-up of LCOE. Invest = investment costs., New stack SOFC contains the SOFC investments that have to be done during the project lifetime.

The investment costs make up the largest part of the LCOE as can be seen in Graph 2: 80% including later SOFC stack replacements and 65% without later stack replacements. The largest investment is the SOFC system, which makes up 27% of the LCOE only by initial investment, and even 42% with both initial investment and later stack replacements.

To investigate the economics with less influence of investments in SOFC and other components, the NPV and LCOE were calculated for three pairs of scenarios. In one pair the SOFC costs have decreased because of learning, in another pair all initial investments have been removed from the analysis and in a final pair only part of the system is built followed by incineration. All results are summarized in Graph 3. The main conclusions are that SOFC learning alone will not make the system feasible, even when subsidy is included. A system with only OPEX costs will be profitable, but not when later stack replacements for SOFC are included. If only part of the installation would be build and combined with the incinerator at the AEB site, the most profitable option would be to only dry the WTP to 25% moisture. Thus, the original system can only become feasible when all initial investments including stack replacements are considered as sunk costs. Though, the most profitable way to convert the waste toilet paper is to just dry and incinerate it.
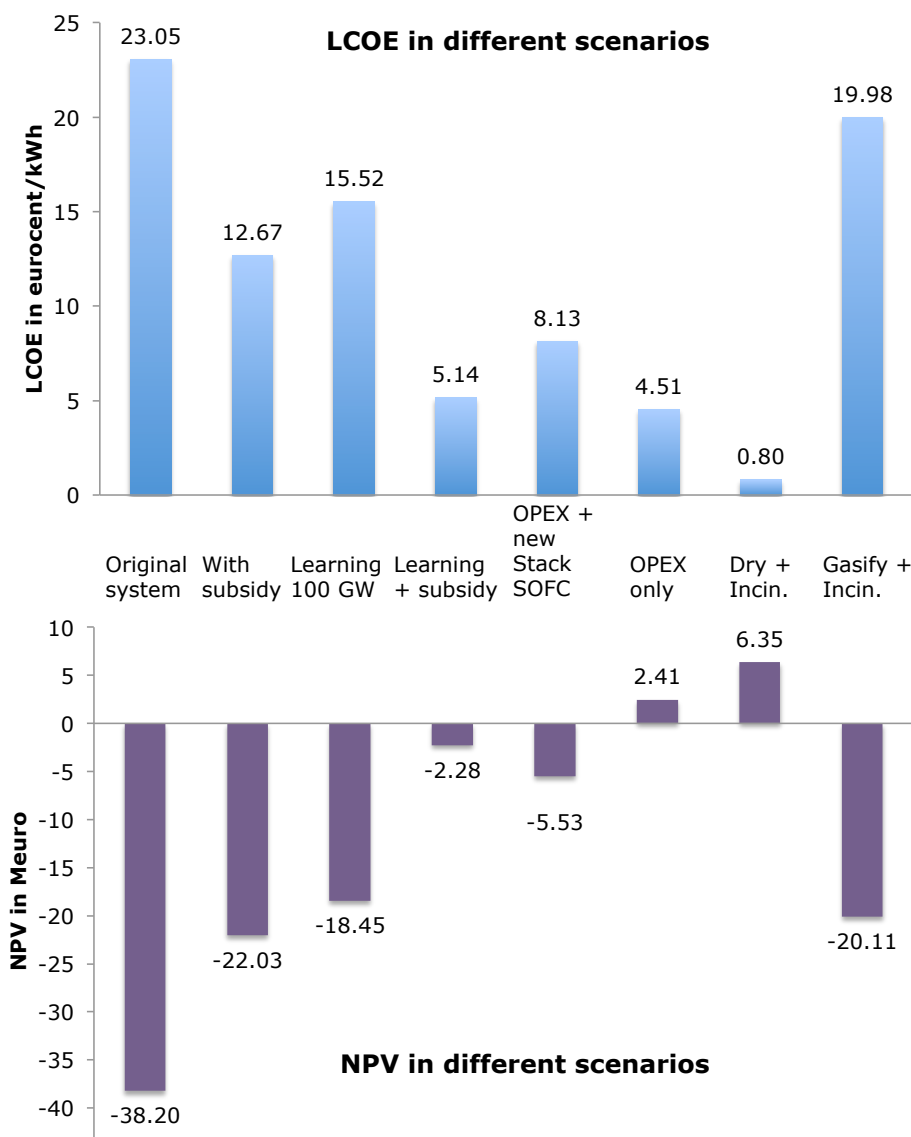
Graph 3 LCOE (above) and NPV (below) for different scenarios from left to right: The original system and system with subsidy; Then the system when the SOFC market has grown to 100GW, with and without subsidy; Next two systems with only OPEX and new SOFC stacks or OPEX only; Then finally two incineration scenarios, one with only WTP drying followed by incineration, the other with drying, gasification and cleaning followed by incineration.

Uncertainty

Thirteen input parameters were varied over an appropriate range to evaluate their impact on the LCOE and combined with the scores obtained from the pedigree analysis to create a diagnostic diagram (Graph 4).This diagram helps to identify the largest sources of uncertainty, especially by pointing out the parameters with low strength (pedigree) but high impact on the LCOE. These parameters are SOFC investment cost and water content of WTP, and to a lesser extent the amount of WTP.
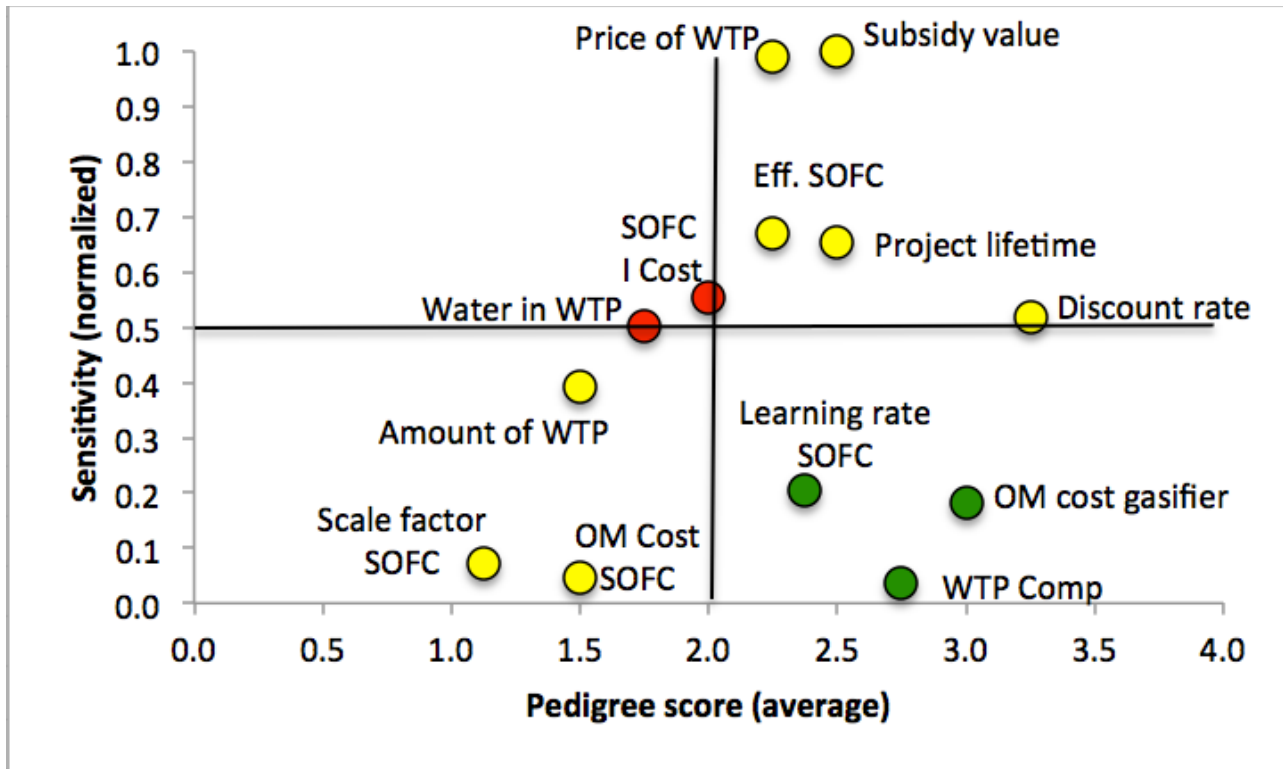
Graph 4 Diagnostic diagram. The x-axes shows the strength of the parameters as average of the pedigree score. The y-axes shows the normalized sensitivity of the parameters on the LCOE.

Discussion of the methodology

One of the main assumptions in this study is that the mechanisms behind wood gasification and gasification of WTP are the same. This assumption was made in order to create an excel model for the MILENA gasifier based on ECN data from wood. This assumption causes a large source of uncertainty in the product gas composition that could not be dealt with during the sensitivity analysis. It is expected that gasification of WTP will give different results compared to wood due to the higher cellulose content, but this was not verified with experiments and therefore causes an unknown amount of uncertainty in the results.

Another point of discussion is the heat integration. This was done in a simplified way by essentially just matching heat sources and sinks. For a more accurate idea of the heat integration possibilities a proper pinch analysis is required. Furthermore, during the economic analysis, the SOFC costs had to be based on literature data from general market reports. The economic analysis would be more accurate when quotations from SOFC suppliers were used.

The pedigree analysis could have been done more extensively, with more people and more experts on (certain parts of) the research. Now some people did have experience with pedigree analysis, but not with all parts of the research, while others could be considered as experts on certain parts, but were not familiar with pedigree analysis. Furthermore, the information document used during the pedigree analysis could be improved in order to be more neutral and complete.

Conclusion and Recommendations for AEB/Waternet

The study showed that it is certainly possible to build a 5MWth installation that converts WTP into 2.8MW electricity with an indirect gasifier and SOFC. However, the system is not feasible at the moment with an LCOE of 23.05 eurocent/kWh and an NPV of -38.20 Meuro. This could decrease to 15 eurocent/kWh with a net present value of -18.45 Meuro when the SOFC market grows to 100GW cumulative global shipments. In both cases the system does not reach break even, which makes it a bad investment from an economic perspective. Subsidy could help, but with a subsidy of 10.38 eurocent/kWh as analysed in this research, the NPV is still -2.3 million euros. Only when the economic analysis is solely based on OPEX cost, the system becomes profitable at an LCOE of 4.51 eurocent/kWh with an NPV of 2.41 million euro.

An alternative option could be to only dry the WTP to 25% moisture and then incinerate it at the AEB with an LCOE of 0.802 eurocent/kWh and an NPV of 6.35 million euro. In this case, the electricity yield will be 1.8 MW. Although the incineration option is more feasible, the electric efficiency of the gasifier-SOFC system is 57.2%, which makes it a promising option for high efficient electricity production (from waste) in the - near - future. Therefore, the recommendations based on the study are:

- Closely watch the SOFC price development, the technology is now in an early commercial phase and learning will lead to reduced system cost. However, this will not reduce the cost enough to make the system feasible, so investigating more subsidy options is recommended as well.
- Investigate ways to increase the scale of the project by combining different waste (cellulose) streams or combining WTP from different locations. This will increase the feasibility of the project.
- When in a later stage a more detailed feasibility study is done, parameters that need most attention are:
  - o SOFC investment costs: The costs are currently based on open source data, mainly from fuel cell market reports. It would be more accurate to obtain direct quotations from SOFC suppliers at the required scale.
  - o Water content and amount WTP: These two parameters are interrelated. The 60% water content is estimation from Waternet, but not confirmed with a reasonable amount of experiments. By pressing the WTP, the water content at the gate of the AEB can be confined to 60%. Though, this could affect the amount of waste toilet paper available. It is therefore advised to investigate what will be the exact amount of waste toilet paper available at fixed moisture content.

1. Introduction

Toilet paper is an essential product for most people in the Western world. Probably, most people cannot imagine a life without it, though at the same time they will not realize or even think about the impact of using all this paper. An average European person consumes 10-14 kg of toilet paper each year (Ruiken et al., 2010). In the Netherlands used toilet paper is flushed and then processed at a wastewater treatment plant (WWTP), where it generally becomes part of the final waste stream (sludge). However, waste toilet paper (WTP) still contains cellulose, which is valuable material.

The water company Waternet in Amsterdam is trying to separate waste toilet paper from the sludge. This could have a positive effect on their total energy balance as less energy is needed for sludge processing. However, Waternet currently has to pay for the external processing of the WTP, which makes the separation process less attractive. In theory it would even be possible to produce new paper out of WTP. Unfortunately the packaging company Smurfit Kappa and 'kenniscentrum papier en karton' concluded that this type of paper will not be accepted by consumers because of hygienic and image issues (Ruiken et al., 2010). Therefore, the challenge is to find another way to add value to 25.000 tonnes[1] waste toilet paper (with 60% moisture) per year.

STOWA (Stichting Toegepast Onderzoek Waterbeheer) published a report in 2010 (Ruiken et al., 2010) that discusses different ways of valorizing this waste toilet paper or 'zeefgoed' (WTP). The possibilities include fatty acid production, recycling to new paper and three different ways of fermentation. The options for fermenting were to ferment the waste toilet paper with the sludge, separate fermentation to biofuel/biogas, or partial fermentation to fertilizer. At the moment none of these options are implemented on a large scale.

In this research another route is proposed: converting the waste toilet paper to electricity. In this way, a waste steam with a more or less continuous supply is converted into a product with continuous demand. The advantage of this route is that local electricity production from a waste source decreases our dependence on fossil fuels and can increase the security and sustainability of energy supply (Mcphail, Cigolotti, & Moreno, 2012). Besides, there is no food or land competition, which is often a problem with other sources of biomass (Mcphail et al., 2012). Because WTP is made from trees that are part of the short carbon cycle, the greenhouse gas emissions are minimal when compared to fossil fuels.
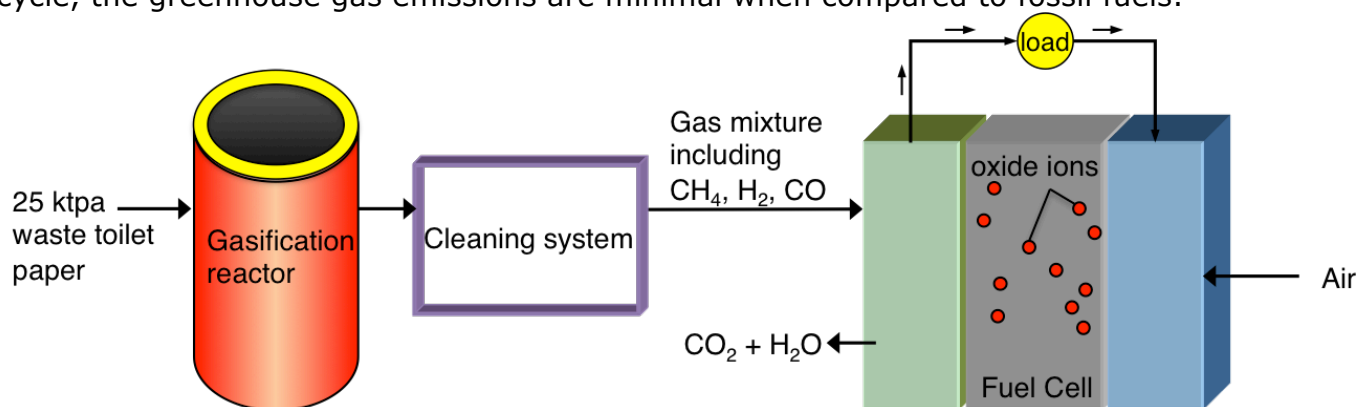
Figure 3 Illustration of the two-step process for converting waste toilet paper to clean electricity.

The WTP to electricity route proposed in this study consists of two main steps, which are gasification of the waste toilet paper followed by utilization of the product gases in a fuel cell to generate electricity, see Figure 3 for an illustration. The first calculations show that about 6400 average Dutch households could be provided with electricity from WTP at the WWTP in Amsterdam-West. The underlying calculations are summarized in Appendix A.

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[1] Amount of WTP from the WWTP in Amsterdam-West per year, based on personal communication (C. Reijken, Waternet) and Ruiken et al. (2010).

1.1. Research goal

The concept of WTP to electricity and the first calculations triggered interest from the Afvalenergiebedrijf (AEB) Amsterdam, a company that is specialized in producing energy from waste. The AEB may want to invest in a pilot plant, but first the management needs more information about the energy balance and economic feasibility of the process. Thus, the following research goal was formulated:

The goal of this study is assessing the techno-economic feasibility of a waste toilet paper to electricity plant.

The following questions will be answered to contribute to this goal:

5. *What could be a possible process design for a WTP to electricity plant?*
   The drawing in Figure 3 is a very simple representation of the most essential parts of equipment of the process. For a feasibility study, it is necessary to identify possible other process steps and to assess how they could be coupled into a functioning WTP to electricity plant.
6. *How would such process perform in terms of energy efficiency and electricity yield?*
   To answer this question, the chemical and thermal energy inputs and outputs of every sub-process in the scenario have to be determined, including mass flows. Also, possibilities of heat integration have to be considered.
7. *What is the economic feasibility of the process?*
   An economic analysis will be performed regarding all the costs and benefits for the plant to determine if building a WTP to electricity plant is worth the investment. Economic indicators such as the payback period (PBP), the net present value (NPV), internal rate of return (IRR) and levelized cost of electricity (LCOE) will be calculated. Also learning curve analysis will be done to analyse how the LCOE could develop in the future.
8. *What are the largest sources of uncertainty in the results?*
   Every investment includes a certain risk. Especially with new process set ups information about the largest risk inducing parameters is very valuable. An uncertainty analysis allows indicating factors or parameters that are not robust and have a large impact on the results; this gives a clear focus for further research.

These questions will be answered for a scenario which consist as much as possible of technologies that are commercially available. The results may lead to a decision regarding the development of a waste toilet paper to electricity plant, thereby turning a currently useless product into something valuable. From a scientific point of view, the results of this study will contribute to existing literature by presenting a design for a new waste to electricity route. Besides, the economic results give more insight in the feasibility of relatively small waste/biomass to energy projects.

1.2. Report structure

The report is organized as follows: Section 2 gives a technical system description in which the basic theory of gasification, fuel cells and cleaning methods is presented. Section 3 discusses the methods used to answer the research questions. Section 4 presents the results of the feasibility study. Section 5 contains a discussion of the results as well as the methodology. In the final section (6) conclusions are drawn and recommendations for further research are given.

2. Technical system description

Gasification reactors, fuel cells and cleaning systems are essential parts of the biomass to electricity process (Caliandro, Tock, Ensinas, & Marechal, 2014; van der Meijden, Rabou, Vreugdenhil, & Smit, 2011; Nagel, 2008; Nanou, 2013; van Paasen, Cieplik, & Phokawat, 2006; van der Spek, 2009). Therefore the basics of these technologies are explained here. Section 2.1 explains more about the process of gasification in general and more specifically for medium and high temperature gasification. Section 2.2 discusses different options for gas cleaning in order to use the gas in a fuel cell. Section 2.3 elaborates on the concept of a fuel cell and describes a fuel cell that could be used in the process of converting WTP to electricity.

2.1. Gasification

Gasification comprises the conversion of a carbonaceous feedstock to a mixture of gasses including hydrogen (H_2), carbon monoxide (CO), carbon dioxide (CO_2) and methane ($CH_{4)}$. Gasification typically involves three steps (executed in the same reactor): drying, pyrolysis, and partial oxidation. During pyrolysis, also called devolatization, volatiles such as hydrogen, methane, carbon monoxide and carbon dioxide are released and char is produced (Highman & van der Burgt, 2008). Afterwards, the remaining char reacts at a higher temperature with oxygen, steam or carbon dioxide to form mainly hydrogen and carbon monoxide (Highman & van der Burgt, 2008; Mcphail et al., 2012). Also some tars are formed, which are a mixture of (polyaromatic) hydrocarbons with a molecular weight higher than benzene (Aravind & de Jong, 2012; van Paasen et al., 2006).

Thus, instead of complete combustion to water and carbon dioxide, the reactants are partially oxidized due to a low oxygen-fuel ratio (Phillips, 2006). The overall reaction is exothermic, which means that the (partial) burning of char provides enough heat for the endothermic gasification reactions. Nanou (2013) gives a general reaction (R1), though in fact many different reactions are happening at the same time, which is illustrated in Table 1.

$$C_xH_yO_z + aH_2O \; or \; bO_2 \rightarrow cCO \; + \; dH_2 + \; e\,CO_2 + fCH_4 + gC_{2-4} + hTars + iC(s) \qquad \text{(R 1)}$$
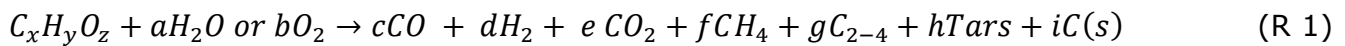
By altering the feed of steam or oxygen/air, the pressure, and the temperature, a different ratio of products can be obtained (Highman & van der Burgt, 2008; van der Meijden, Rabou, Vreugdenhil, & Smit, 2011; van der Meijden, 2010; Nanou, 2013; Phillips, n.d.). Temperature is generally selected based on the ash properties, because the reactor will have a high fouling rate when the ash is between its softening and melting point (Highman & van der Burgt, 2008). The two main types of gasification are described in more detail below.

Table 1 Main gasification reactions at 25°C, taken from (Ruiz, Juárez, Morales, Muñoz, & Mendívil, 2013)

	Reaction	dH (kJ/mol)
Char or gasification reactions		
1 (Boudouard)	$C + CO_2 \leftrightarrow 2CO$	+172
2 (Steam reforming)	$C + H_2O \leftrightarrow CO + H_2$	+131
3 (Hydrogasification)	$C + 2H_2 \leftrightarrow CH_4$	- 74.8
4	$C + 0.5\,O_2 \leftrightarrow CO$	- 111
Oxidation reactions (combustion)		
5	$C + O_2 \leftrightarrow CO_2$	- 394
6	$CO + 0.5\,O_2 \leftrightarrow CO_2$	- 284
7	$CH_4 + 2\,O_2 \leftrightarrow CO_2 + H_2O$	- 803
8	$H_2 + 0.5\,O_2 \leftrightarrow H_2O$	- 242
Water gas shift reaction		
9	$CO + H_2O \leftrightarrow CO_2 + H_2$	- 41.2
Methanization reactions		
10	$2CO + 2\,H_2 \leftrightarrow CH_4 + CO_2$	- 247
11	$CO + 3\,H_2 \leftrightarrow CH_4 + H_2O$	- 206
12	$CO_2 + 4\,H_2 \leftrightarrow CH_4 + 2\,H_2O$	- 156
Steam reactions		
13	$CH_4 + H_2O \leftrightarrow CO + 3H_2$	+ 206
14	$CH_4 + 0.5\,O_2 \leftrightarrow CO + 2H_2$	- 36

2.1.1. High temperature gasification

When the desired gasification product is pure syngas[2], the gasification reaction is typically performed at high temperature to prevent formation of methane and tars (1300– 1500°C) (Van der Meijden, 2010). For high temperature gasification, mostly entrained flow gasifiers are used (Figure 4). Entrained flow gasifiers have a feed of finely ground fuel and air or steam from the top (or sides), need high temperatures and the reactants have a short residence time (Nanou, 2013; Phillips, 2006). Operation is complex and it requires high oxygen consumption and low variability in the feedstock, but this type of gasifier is easy to scale up (Highman & van der Burgt, 2008). Although the gasifier and temperature conditions are favourable for syngas production, the product will not be clean syngas. After gasification cleaning steps will be required to filter out contaminant gases such as H_2S, NH_3 or HCl that arise from the S, N and Cl components in the biomass (Aravind & de Jong, 2012; van der Meijden, Veringa, Vreugdenhil, & Drift, 2009; Nanou, 2013).
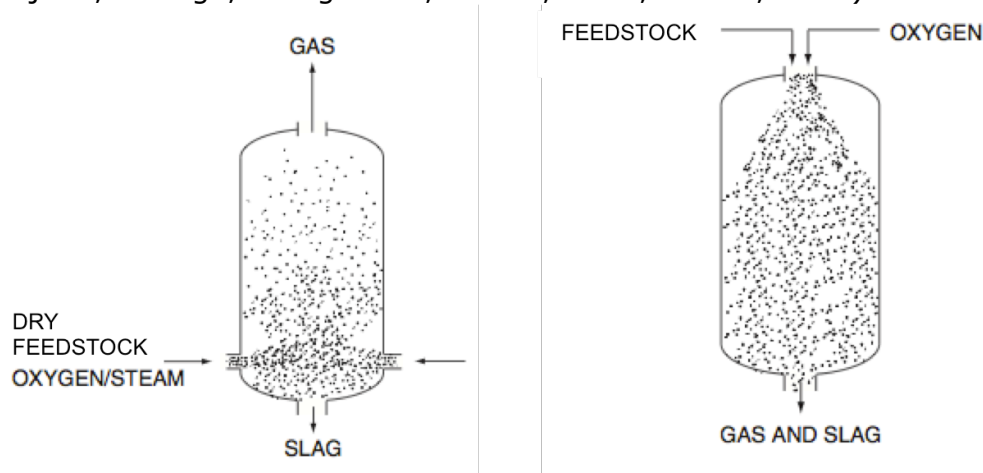
Figure 4 Entrained flow gasifiers, feeded from the sides (L) or from the top (R). Based on (Highman & van der Burgt, 2008).

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[2] An almost pure mixture of hydrogen (H_2) and carbon monoxide (CO).

2.1.2. Medium Temperature gasification

Medium temperature gasification results in a mixture that contains the useful product gasses H_2, CO and CH_4, also called producer gas. The gasification temperature is 700-900 °C (van der Meijden et al., 2011; Nanou, 2013). For this kind of gasification, two main types of gasifiers exist, the fixed/moving bed gasifier and the fluidized bed gasifier, visualized in Figure 5.

The fixed/moving bed gasifier has low investment costs and a simple structure. However, for this gasifier a constant and homogeneous feedstock size is required and upscaling is difficult (van der Meijden et al., 2011; Ruiz et al., 2013). Solid fuel enters the reactor from the top while air or steam enters from below and moves up, thereby mixing with the fuel. Producer gas leaves at the top (updraft) or at the bottom (downdraft). A disadvantage is that ashes and fines can block the gas from flowing (Higman & van der Burgt, 2008).

The fluidized bed gasifier has a higher oxygen flow or air flow and the feedstock is more uniformly mixed with the fuel than in a fixed/moving bed gasifier. The residence time of the reactants is typically lower than in a fixed/moving bed gasifier, but the operation is somewhat more complex (Philips, n.d.). Besides, the conversion efficiency is lower because some unreacted fuel will be part of the ash due to the mixing (Higman & van der Burgt, 2008).

A distinction can be made between different types of fluidized bed gasifiers. Figure 5 shows the bubbling fluidized bed (BFB), the circulating fluidized bed (CFB) and the indirect fluidized bed gasifier[3]. The BFB and CFB work with one vessel where both gasification and combustion are happening at the same time, just as in the fixed/moving bed gasifier. The indirect gasifier however works with two separate rooms for combustion and gasification, and in this way the combustor can utilize air as oxidant without the drawback of diluting the producer gas with large amounts of N_2. Furthermore, this type of gasifier can handle a broader range of feedstock (van der Meijden et al., 2011). Similar to high temperature gasifiers, medium temperature gasifiers produce impurities like H_2S and tars that have to be removed before the gas can be used.
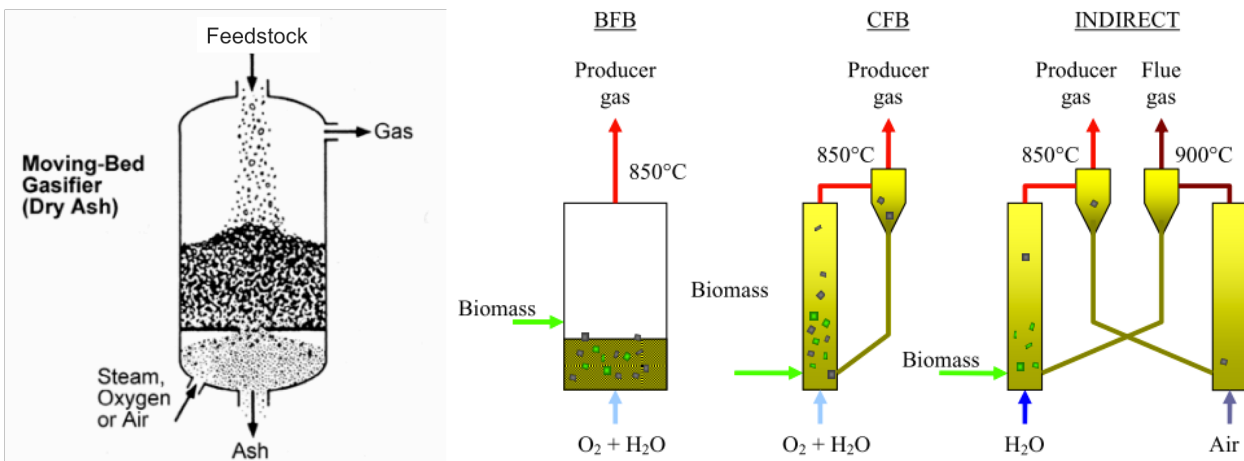
Figure 5 Left: Fixed/Moving bed in updraft configuration based on (Philips, n.d.). Right: Illustration of 3 types of fluidized bed gasifiers: the Bubbling fluidized bed (BFB), circulating fluidized bed (CFB) and indirect fluidized bed. Illustration from (van der Meijden et al., 2011).

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[3] Also called dual or allothermal

2.2. Fuel Cell

Fuel cells are very efficient converters of chemical energy to electrical energy. Because no intermediate step is required in which fuel is converted into heat, fuel cells are not hindered by the Carnot efficiency. Their electrical efficiency can theoretically go up to 80% (da Rosa, 2009). Currently, commercial fuel cells have a practical electrical efficiency of around 50-60% (Ammermann et al., 2015; James, Spisak, & Colella, 2012; Mcphail et al., 2012), with a maximum of 70% if a gas turbine is added to convert unreacted fuel (McPhail et al., 2012). Possible fuels include hydrogen, syngas and methanol, but these fuels typically need to be very clean, i.e. virtually free of contaminants and impurities (Aravind & de Jong, 2012; Caliandro et al., 2014; Mcphail et al., 2012). Table 2 presents an overview of different types of fuel cells and their main characteristics

Table 2 Fuel cell types and main characteristics.

	Fuel	Operating Temperature (°C)	Electrolyte	Electric efficiency (%)	Applications
Alkaline Fuel Cells (AFC)	H_2	60-120	KOH	45-70	Military Spacecraft
Proton Exchange Membrane Fuel Cell (PEMFC)	H_2 (/CO_2)	40-100	Polymer	30- 60	Transport, CHP, mobile and stationairy applications
Direct Methanol Fuel Cell (DMFC)	Methanol (CH_3OH)	60-130	Polymer	20-30	Mobile applications, transport
Phosphoric Acid Fuel Cell (PAFC)	H_2 (/CO_2)	160-220	Phosphoric acid	35-50	Decentral power, CHP
Molten Carbonate Fuel Cell (MCFC)	H_2 ,CO	600- 650	Molten carbonate	45-60	Decentral power, CHP
Solid Oxide Fuel Cell (SOFC)	H_2 ,CO ,small hydrocarbons (CH_4)	650-1000	Solid oxide	45-60	Decentral power, CHP

Sources (Bocci, Sisinni, et al., 2014; Nagel, 2008; Nieuwlaar et al., 2014)

The working mechanism is the same for all fuel cell types and is similar to a battery system. The difference lies in the fact that batteries are closed systems while fuels cells are open systems that are continuously fuelled with gasses at the electrodes (Ellamla, Staffell, Bujlo, Pollet, & Pasupathi, 2015; da Rosa, 2009). The principle of reduction and oxidation reactions at the anode and the cathode is the same; electricity is through separation of the two half-reactions.

This research focuses on the stationary application of a fuel cell that has to deal with the main gasification products of a biomass feedstock, i.e. CH_4, H_2, CO, and CO_2. For this purpose, often a Solid Oxide Fuel Cell (SOFC) is used (Bocci, Di Carlo, et al., 2014; Broust, 2008; Nagel, 2008). Therefore, the SOFC is described in more detail below.

2.2.1. Working mechanism SOFC

A normal SOFC utilizes H_2 and CO at the anode and O_2 (pure or from air) at the cathode according to the half reactions R2-R4. Moreover, an SOFC is capable of the internal reforming of some CH_4 followed by the oxidation of the formed H_2 and CO (James et al., 2012). The total reaction of reforming and oxidation is given in R5.

<u>Cathode</u>	$\frac{1}{2}O_2 + 2e^- \rightarrow O^{2-}$	(R 2)
<u>Anode</u>	$H_2 + O^{2-} \rightarrow H_2O + 2e^-$	(R 3)
	$CO + O^{2-} \rightarrow CO_2 + 2e^-$	(R 4)
	$CH_4 + 4O^{2-} \rightarrow 2H_2O + CO_2 + 8e^-$	(R 5)
Total reaction		
1)	$\frac{1}{2}O_2 + H_2 \rightarrow H_2O$	(R 6)
2)	$\frac{1}{2}O_2 + CO \rightarrow CO_2$	(R 7)
3)	$2O_2 + CH_4 \rightarrow CO_2 + 2H_2O$	(R 8)

The SOFC electrolyte is ceramic and can withstand its high operating temperature. The most commonly used electrolyte is yttria stabilized zirconia (YSZ or $(ZrO_2)_{0.9}(Y_2O_3)_{0.1}$) (da Rosa, 2009). This electrolyte has as characteristic that it is conductive for ions, but not for electrons (Ellamla et al., 2015; Mahato, Banerjee, Gupta, Omar, & Balani, 2015; Nagel, 2008). A popular material for the electrode is strontium- doped lanthanum manganite, LSM, while the anode is mostly a combination of YSZ with nickel which is called a cermet (da Rosa, 2009). Figure 6 illustrates the working principle: it shows that the reduced oxygen ion O^{2-} travels through the electrolyte and reacts with the fuel at the anode side. So the air that leaves the fuel cell at the cathode side is leaner in oxygen content, while the outlet at the anode side contains some unreacted fuel plus CO_2 and H_2O.
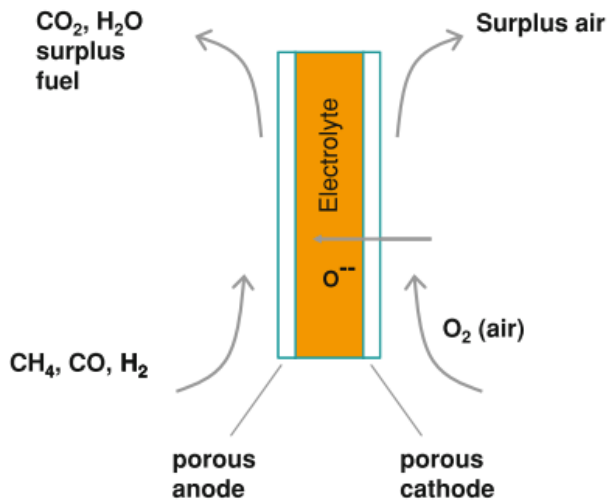
Figure 6 Simple representation of an SOFC, picture from (Mcphail et al., 2012).

Because the operating temperature of an SOFC is high, it is good for Combined Heat and Power (CHP) applications (Ellamla et al., 2015; Da Rosa, 2009). Besides, CHP application of an SOFC leads to a high overall efficiency of energy conversion.

An SOFC can internally reform and convert CH_4 due to catalytic activity of nickel in the anode. However, a main risk of this reaction is carbon deposition on nickel (Figure 7), leading to decreased performance. This can be avoided by using large amounts of steam or using a different type of catalyst (Fuerte, Valenzuela, Escudero, & Daza, 2014; Klein, Georges, & Bultel, 2010; Laosiripojana & Assabumrungrat, 2007; Meng et al., 2014). In general, these changes lead to a decrease of electrochemical efficiency and lifetime.
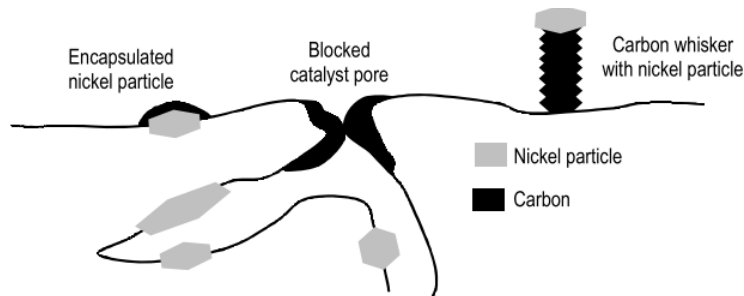
Figure 7 Carbon deposition on the nickel catalyst of the SOFC, from (Seemann, 2007)

On the other hand, many commercial fuel cells are fed with natural gas that consists mainly of CH_4. As not all CH_4 can be internally reformed due to abovementioned problems, in general a pre-reformer is part of the system and converts part of the methane to pure syngas via reaction 8 (Ammermann et al., 2015; James et al., 2012; Leo, n.d.). According to James et al. (2012), about 25% of the gas has to be pre-reformed before entering the fuel cell. The heat required for this steam-reforming reaction is obtained from the exothermic fuel cell process.
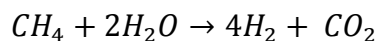
$$CH_4 + 2H_2O \rightarrow 4H_2 + CO_2 \hspace{4cm} \text{(R 9)}$$

2.3. Cleaning equipment

The cleaning equipment forms the connection between the gasifier and the fuel cell and should reduce the contaminants in the producer gas to fuel cell inlet requirements. These requirements include the virtual absence of sulphur and chlorine containing compounds, as well as absence of particulates and tar. Producer gas typically contains all these impurities. In this section, I will discuss different contaminants, explain the problems related to the contaminant and discuss possible cleaning mechanisms.

2.3.1. Particulates and dust

Particulates are all kinds of small particles in the gas that are not in gas phase such as ash and char. The exact effect of these particles on SOFC and thus the tolerance levels are unknown (Aravind & de Jong, 2012). Still one wants to avoid ash and char reaching the SOFC, for example to prevent clogging of the anode pores. The most common ways for char and ash removal from the gasifier are through a cyclone or candle filter (Highman & van der Burgt, 2008; Phillips, n.d.; Zwart, 2009), see Figure 8. A cyclone uses centrifugal forces to separate small particles from the gas (Zwart, Boerrigter, Deurwaarder, Meijden, & van Paasen, 2006) while a candle filter uses filter candles (Highman & van der Burgt, 2008).
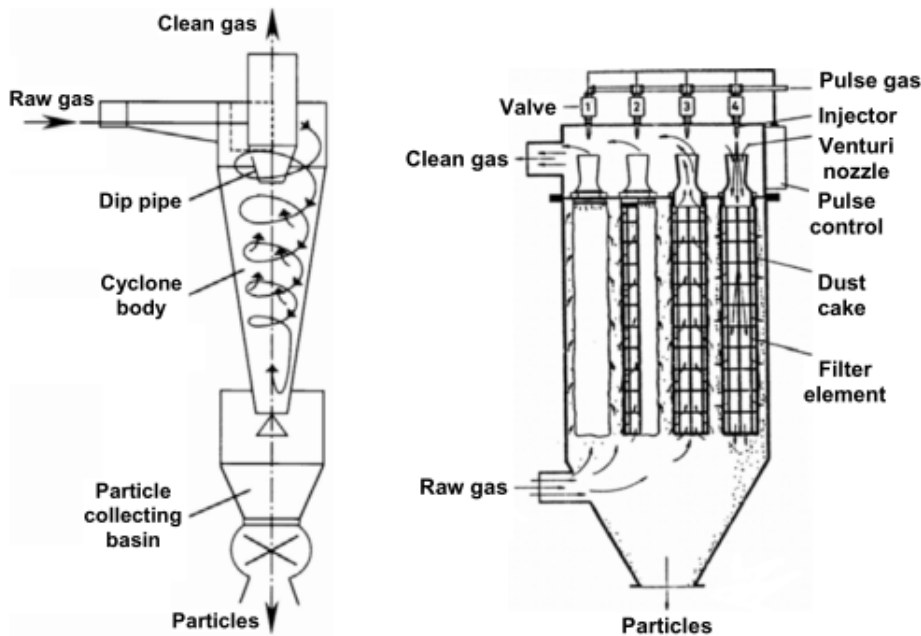
Figure 8 Cyclone (left) & candle filter (right), source (Nagel, 2008)

2.3.2. Tar

Tars are a mixture of (polyaromatic)hydrocarbons with a molecular weight higher than benzene (Aravind & de Jong, 2012; van Paasen et al., 2006). For the SOFC, it is not completely sure if the tars are a problem, because it could be that tars are internally reformed by the nickel in the anode, just as methane. Nagel (2008) stated that tars could in thereby even be a fuel for SOFCs. However, tar is usually considered as a problem, because it deposits on the wall of equipment if the gas cools down below the tar dew point[4], causing fouling of the system. Options for tar removal are for example ECN's OLGA tar removal system, catalytic tar cracking, or catalytic partial oxidation.

The OLGA technology is able to recycle all tars in order to use them as fuel in the gasifier system (Boerrigter et al., 2005). This reduces waste and increases the energy efficiency of the entire system. The tar dew point can be decreased to <20°C degrees with OLGA (Zwart, 2009). The technology works with three stages, (Figure 9). First, the heavy tars are removed by scrubbing oil during cooling of the gas from 400 °C to about 80 °C in the collector. The tars are recycled to the combustor where they are burned, together with a small bleed of the scrubbing oil (0.1 g/Nm³ gas) (Boerrigter et al., 2005). The second step is absorption where light tars are absorbed by the scrubbing oil. Finally, in the stripper, the light tars are released from the oil with hot air that can be recycled to the gasifier. The mass flow of air can be adapted to the airflow that is used in the gasifier. In that way, the air flows directly from OLGA to the combustor without the need for extra air addition (Boerrigter et al., 2005).

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[4] Which is typically arount 350°C when coming out of the gasifier (Boerrigter et al., 2005)
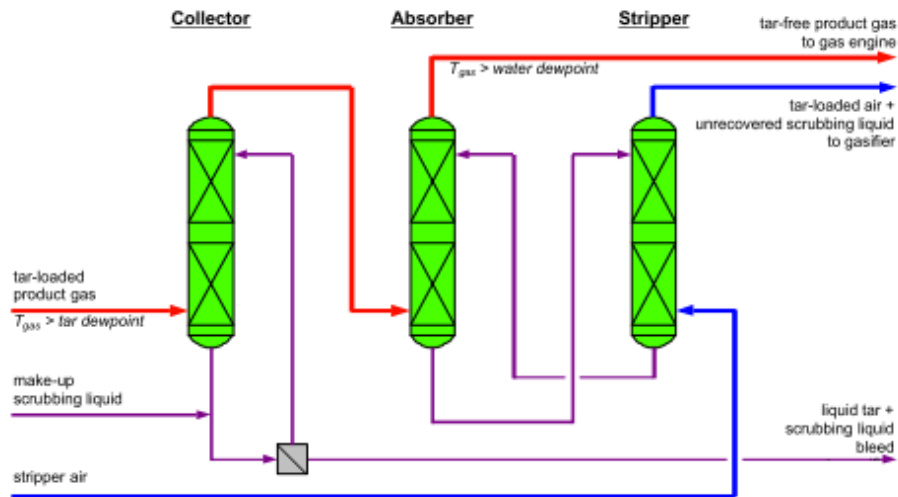
Figure 9 Outline of the OLGA tar removal system (Boerrigter et al., 2005).

Catalytic tar removal involves cracking the tars in a catalytic filter. As catalysts calcined carbonate rocks or nickel-based catalysts can be used (Aravind & de Jong, 2012). Using nickel will result in better tar conversion (Aravind & de Jong, 2012), but the drawback is that nickel is deactivated by carbon deposition as was explained in 2.2.1. Therefore, another option could be to use two stage tar removal with calcined carbonate rocks (dolomite) in the first stage and nickel in the second stage (Caliandro et al., 2014).

The last tar removal option discussed here is tar conversion by catalytic partial oxidation (CPO), tested by Nagel (2008). A CPO reactor consists of ceramic monoliths coated with noble metals. An extra advantage here is that organic sulphur species (see 2.3.4) are converted as well. Though, the system is still in development and was only tested on 1 kW scale (Nagel, 2008).

2.3.3. HCl

Hydrogen chloride (HCl) gas should be removed before feeding the gas to an SOFC because it can cause corrosion to system components. Besides, it could react with components on the anode which will cause degradation (Aravind & de Jong, 2012). To remove HCl, both wet and dry srubbing methods could be used, though dry scrubbing is cheaper and works on higher temperature, thereby reducing energy losses that occur while cooling (Zwart et al., 2006). For dry srubbing typical sorbents of good quality are sodium bicarbonate ($NaHCO_3$) or sodium carbonate (Na_2CO_3) that are both able to remove HCl to < 1 ppmv and can work around temperatures up to 600°C (Aravind & de Jong, 2012). The reactions during the HCl removal are (Aravind & de Jong, 2012):

$$NaHCO_3 \ (s) + HCl \ (g) \rightarrow NaCl \ (s) + H_2O \ (g) + CO_2 \ (g) \hspace{2cm} \text{(R 10)}$$
$$Na_2CO_3 \ (s) + 2HCl \ (g) \rightarrow 2NaCl \ (s) + H_2O \ (g) + CO_2 \ (g) \hspace{1.5cm} \text{(R 11)}$$
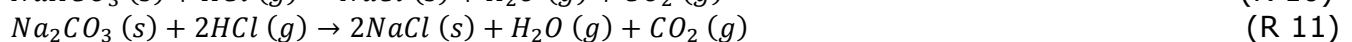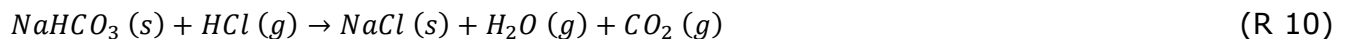
For wet scrubbing of HCl, a water scrubber can be used. Water (just below 100°C) will let the HCl dissolve and the HCl then leaves the system with the scrubbing water, as well as some trace particulates and ammonia (van der Spek, 2009).

2.3.4. Sulphur species

The main sulphur species in producer gas from the gasifier are hydrogen sulphide (H_2S), COS and thiophene (C_4H_4S) (van der Meijden, 2010). Though biomass in general has a lower sulphur content than coal, fuel cells are very sensitive to sulphur poisoning of the anode (Aravind & de Jong, 2012). Thus, all kinds of sulphur should be removed as much as possible (Aravind & de Jong, 2012; Bocci, Di Carlo, et al., 2014; Mcphail et al., 2012).

Thiophene removal is usually done in an hydrodesulphurization reactor (HDS) with the help of an Cobalt-Molybdenum (Co-Mo) or Nickel-Molymbdenum (Ni-Mo) catalyst (Highman & van der Burgt, 2008; Rabou & Almansa Aranda, 2015; Zuber, Hochenauer, & Kienberger,

2014; Zwart, 2009). The catalyst is only active above 350°C and during the process exothermic reactions occur which means that the outlet temperature is between 500°C-550°C (Highman & van der Burgt, 2008; Zuber et al., 2014).

A 500 hour test from ECN with an HDS reactor showed that next to 100% thiophene conversion COS is completely converted in the reactor as well (Rabou & Almansa Aranda, 2015). Furthermore, about 98% HCN removal was reached and hydrocarbons with double bonds were converted to single bonds by hydrogenation (Rabou & Almansa Aranda, 2015). To summarize, the reactions that can happen in the HDS are given in R12-R15. Obviously, this reactor should be placed in front of a potential H_2S removal system, as more H_2S is produced during the conversion of COS and thiophene.

Thiophene	$C_4H_4S + 4H_2 \rightarrow C_4H_{10} + H_2S$	(R 12)
COS	$COS + H_2 \rightarrow CO + H_2S$	(R 13)
HCN	$HCN + 3H_2 \rightarrow CH_4 + NH_3$	(R 14)
Hydrocarbons (i.e. ethylene)	$C_2H_4 + H_2 \rightarrow C_2H_6$	(R 15)

When thiophene and COS are converted, the H_2S should be removed from the gas to sub-ppm levels. H_2S is a very poisonous compound, not only for fuel cells but for humans as well (Bovenkamp, 2009). Hence, many methods for H_2S removal from gases have been developed, varying from dry adsorption via absorption in alkali solution, liquid oxidation and molecular sieves to biological removal (Mcphail et al., 2012; van Paasen et al., 2006; Zwart, 2009).

Because of the relatively small scale (around 5 MWth producer gas, see appendix A), dry adsorbents and absorbents are suitable. Metal oxides are a very good option for deep H_2S removal and are able to work at elevated temperatures (Aravind & de Jong, 2012), which would be practical when the system is placed behind the HDS reactor. For temperatures above 500°C, Ceria based sorbents could be a good option (Aravind & de Jong, 2012). Under 500°C, removal of H_2S from producer gas after gasification by means of zinc oxide is common (van der Meijden, 2010; van Paasen et al., 2006; Rabou & Almansa Aranda, 2015; Zuber et al., 2014; Zwart et al., 2006) also when the gas is used for fuel cells later on (Aravind & de Jong, 2012; Bocci, Di Carlo, et al., 2014; Broust, 2008; Nagel, 2008). ZnO is able to remove H_2S under 1 ppmv concentrations (van Paasen et al., 2006; Zwart, 2009) according to reaction 16.

H_2S	$H_2S + ZnO \rightarrow H_2O + ZnS$	(R16)

2.3.5. Nitrogen species

The nitrogen species in gasification gas mainly comprise ammonia (NH_3) and some hydrogen cyanide (HCN). Ammonia does not seem to be a problem for an SOFC and could even be used as a fuel (Aravind & de Jong, 2012; Bocci et al., 2014; Fuerte, Valenzuela, Escudero, & Daza, 2009; van Paasen et al., 2006). The following fuel cell reaction mechanism is given by Fuerte et al. (2009), that is possibly catalysed by the anode catalyst (nickel):

NH_3 Reforming	$2\,NH_3 \rightarrow N_2 + 3H_2$	(R 17)

The effect of HCN on an SOFC is unclear (Aravind & de Jong, 2012). Some authors consider the removal of HCN (Bocci, Di Carlo, et al., 2014; Nanou, 2013) but do not discuss a specific removal method. However, if an HDS reactor for conversion of thiophene and COS is used, it will also convert 98% of the HCN (Rabou & Almansa Aranda, 2015).

3. Methods

3.1. General demarcation and system outline

Because the goal of this study is to determine the techno-economic feasibility of a WTP to electricity plant, this report looks at the system from the perspective of a potential operator, in this case the AEB. This means that the process only focuses on the process steps under the operators' control, i.e. from the point that the WTP arrives at the plant up to the point that electricity is sold to the electricity grid. Besides, if there is any waste heat, this was seen as an end product for use in the AEB district heating system.

Although the wastewater treatment process is affected when WTP is extracted, the assumption is that this process is already in place. Therefore, the WWTP including WTP separation is outside the scope of this study. For more information on the wastewater treatment process, see appendix B. Furthermore, the feed stream was limited to waste toilet paper from the WWTP at Amsterdam-West. This is was decided together with Waternet for practical reasons, as the WWTP Amsterdam-West covers all waste water treatment of the Amsterdam region and is located directly next to the AEB.

In Figure 10 a general outline of the system is given that includes the gasifier, gas cleaning system and fuel cell. Also, the system boundary is visualized as well as the simplified energy flows. The system will include as much as possible commercially available equipment to get an idea of the currently available options and costs for this system. It also facilitates a more accurate economic analysis later on.
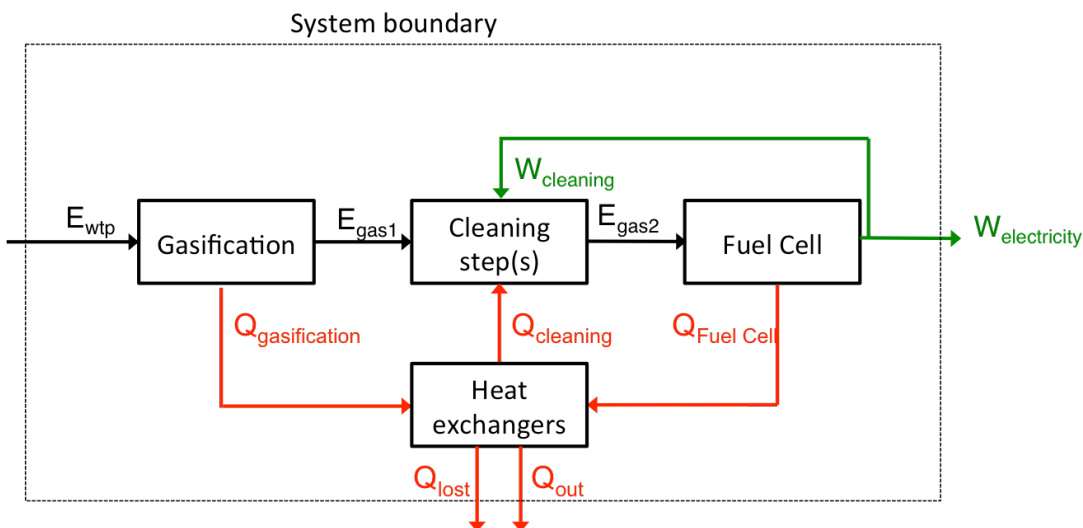
Figure 10 System boundaries and simplified energy flow diagram. E = chemical energy in black, Q = heat flows in red and W is work/electricity flows in green.

3.2. General approach

To conduct a proper early stage techno-economic assessment of a WTP-to-electricity system, the following steps were performed:

1. Fuel characterization
   Because not much was known about the fuel, experiments and calculations were performed to determine the elemental composition and energy content.
2. System design
   Starting from the basic system components (gasifier, cleaning equipment, fuel cell) a more detailed design was created. First, a suitable gasifier was chosen and modelled in excel to find out the producer gas composition. Then an investigation of the fuel cell inlet requirements was performed. Based on the difference between the gasifier output and required fuel cell input, the necessary cleaning equipment was chosen.

3. System analysis
   For every system component, mass and energy balances were created in excel. Subsequently, a simplified heat integration was done to match heat streams. Finally, a total energy balance was created and the energy efficiency and electricity yield of the process were derived.
4. Economic analysis
   Data for the economic analysis were gathered via different sources (literature and equipment suppliers). From the data, some important economic parameters (NPV, IRR, PBP and LCOE) were calculated to give a good indication of the economic feasibility of the system. The possibility of subsidy was investigated as well.
5. Learning effects
   Because both biomass-gasification and SOFC are not (fully) commercialized yet, effects of learning were examined. Learning rates and cumulative installed capacity were derived from literature.
6. Uncertainty
   To check how sensitive the LCOE is for changes in input parameters, a sensitivity analysis was done. A selection of parameters and sensitivity ranges was made. Besides, a pedigree-analysis was performed to check the strength of the values given to the parameters. The results of the sensitivity analysis and the pedigree analysis were combined in a diagnostic diagram.

3.3. Fuel characterization

As waste toilet paper is not a common material for gasification, not much information is available. All currently available information regarding the analysis of contaminants in WTP and the energy content can be found in table 11 of Ruiken et al. (2010). However, more exact knowledge about the elementary composition (also called ultimate analysis) and energy content is required as it is the starting point of many gasification models (Doherty, Reynolds, & Kennedy, 2013; Fryda, Panopoulos, & Kakaras, 2008; Lv et al., 2010; Lv et al., 2007; Nikoo & Mahinpey, 2008; Wu, Wang, Huang, & Williams, 2013).

3.3.1. Ultimate analysis

WTP was obtained from Waternet, from the WWTP at Horstermeer where WTP was filtered out from the primary sludge by means of a sieve. Afterwards it was washed and dried for 2 days at 90°C-95°C. At the UvA, the samples were pulverized with a blender (5 minutes) and 3 samples of 1.5 grams were send to *Mikroanalytisches Labor Kolbe* for ultimate analysis.

3.3.2. Energy content

To calculate energy balance for the system, the energy input is an essential starting point. The energy content of WTP was not measured experimentally in this research, but taken from Ruiken et al. (2010). To check if the number from Ruiken et al. is representative for the material from Horstermeer, the heating value was calculated based on the results from the ultimate analysis. Formulas used were Milne formula for biomass (Eq 1) (Milne, Brennan, & Glenn, 1990) used by ECN (ECN, 2012) and the formula by Channiwala & Parikh (2002) (Eq 2) as given in Nieuwlaar et al. (2014).

Milne:
$$HHV = 0.341 \cdot C + 1.322 \cdot H + 0.0686 \cdot S - 0.12 \cdot O - 0.12 \cdot N - 0.0153 \cdot ash \qquad (Eq\ 1)$$

Channiwala & Parikh:
$$HHV = 0.3419 \cdot C + 1.1783 \cdot H + 0.1005 \cdot S - 0.1034 \cdot O - 0.0151 \cdot N - 0.0211 \cdot ash \qquad (Eq\ 2)$$

Where HHV is the higher heating value in MJ/kg, and C, H, S, O, N and ash are the mass percentages of carbon, hydrogen, sulphur, oxygen, nitrogen and ash on dry basis.

To convert HHV into LHV (Eq 3) formula from (Blok, 2007) was used:

$$E_{LHV} = E_{HHV} - h \cdot E_{w,evap} \cdot m_{H_2O} \hspace{4cm} \text{(Eq 3)}$$

Where E_{LHV} is the lower heating value of the fuel (MJ/kg), E_{HHV} the higher heating value of the fuel (MJ/kg), h the mass fraction of hydrogen in the fuel, $E_{w,evap}$ the heat of evaporation of water at 25°C is 2.442 MJ/kg/K and m_{H2O} the mass of water created per unit mass of hydrogen which is 8.936 kg H_2O/kg H.

3.4. System design

The WTP to electricity system was designed by the following approach: first, a suitable gasifier was selected for gasifying the WTP. Then the composition of the producer gas was calculated, so the exact impurities and amounts were identified. Secondly, a fuel cell (FC) selection was made, including the set of FC performance parameters and gas inlet requirements. Thirdly, given the producer gas composition and the FC inlet requirements, cleaning equipment was selected, leading to a complete system line-up. This design sequence was iterated twice, to arrive at an optimal system design. Once after a discussion with and ECN employee, and once after a discussion with a Royal Dahlman[5] employee. Finally, the system heat flows were integrated, to improve the overall system efficiency.

3.4.1. Gasifier

The first important choice was the type of gasifier that fits the purpose of the project. The gasifier abides by the following requirements:
- Scale of around 5 MWth: this scale was based on preliminary calculations in appendix A.
- Some fuel flexibility: it is not sure yet how homogeneous the WTP will be, plus it would give the AEB a change to try other types of cellulose waste in a later stage.
- Low tar production: tar is a mixed fraction of higher hydrocarbons, its production is unwanted because it is not converted into product gas and can cause fooling of the system.
- Low nitrogen content in the product gas: In the SOFC it would be undesirable to have 80% nitrogen in the inlet stream at the anode. This would require a high circulation rate of inlet gas, which will lead to a decrease in efficiency (Nagel, 2008; van Paasen, Cieplik, & Phokawat, 2006).

To find the right gasifier that could fulfil all requirements, literature research was done (Dascomb & Krothapalli, 2014; Mcphail et al., 2012; van der Meijden, Veringa, & Rabou, 2010; Nagel, 2008; van Paasen et al., 2006; Pfeifer, Koppatz, & Hofbauer, 2011; Ruiz et al., 2013). The paper of Ruiz et al. (2013) includes a very comprehensive table that helped to select gasifiers that could fulfil the first three requirements. Then based on the other literature a gasifier was chosen that could fulfil the last requirement in the most energy efficient way.The gasifier of choice has three possible producers, and the choice between them was made based on the efficiency of the gasification process and the availability of information about the gasifier.

3.4.2. Gas composition after gasification

The basis for the gasification model was an ECN-excel gasifier model (van der Meijden, 2010). At first instance, we planned to use Aspen Plus for modelling the gasifier. However, an interview with Berend Vreugdenhil from ECN[6] made clear that creating a model in Aspen Plus without having any experimental data for verification of the product gas, would give very unreliable results. Therefore, it was chosen to use an excel model baed on the ECN excel model that is partly described by van der Meijden (2010) in his thesis. Two sheets of this exel gasifier model for the gasification of wood (white labee pellets, 25% moisture) are shown in

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[5] Royal Dahlman is the company that sells the MILENA gasifier, OLGA removal system and also necessary cleaning equipment.

[6] See appendix D for interview summaries.

his thesis and are used as the basis for this research (figure 7-19 and 7-20 from van der Meijden, 2010). The assumption was made that the reaction mechanisms are the same, thereby justifying that the ECN model can be used for WTP gasification.

To determine the gas composition after gasification, the ultimate analysis plus all other inlet streams were combined with the experimentally verified gas composition in the model of van der Meijden. The starting point was to convert the mass% of the elements in the ultimate analysis - C, H, O, N, S and Cl- to moles per kg input. The amount of moisture was divided over H and O. The next step was to use the carbon conversion factor to exclude conversion of carbon into char from the product gas. The carbon conversion (CC) can be derived with these formulas from (van der Meijden, 2010):

$C_{conversion} = 79 + 0.04 * (T - 760)$ And is defined as $C_{conversion} = \frac{C_{producer\_gas} - C_{feed-gas} - C_{additives}}{C_{fuel}}$	(Eq 4) (Eq 5)

Where C is the mass flow of carbon in a stream. Carbon in the tar is part of $C_{producer\ gas}$ in this formula.

So the amount of C (in mol/kg feed) was multiplied with the CC. Also the amounts of H and O were corrected according to the ratio of C:H:O in the tar and char. The char and tar composition were taken from van der Meijden (2010) and can be found in Appendix E. Then additional input streams were added as given in the model, such as steam (steam to biomass ratio 0.05), CO_2 and N_2. Furthermore, there is an additional input 1% recycling of gas from the combustor to the gasifier assumed (van der Meijden, 2010). After all these additions, the amount of moles of C,H, N, O, S and Cl per kg of feed were known.

The next step was to determine the ratio in which for instance the element C would be divided over the different products CO, CO_2 and CH_4. We did this by calculating the ratio from the product gas composition (wet) in van der Meijdens model. For instance, when the product gas would contain 18.6 vol% CO and 2.9 vol% C_2H_4 this would be $0.186 * 1 + 0.029 * 2 = 0.244\ moles\ of\ C$. Then the fraction of C in the feed that is converted to CO would be: $\frac{0.186}{0.244} = 0.762$.

Thereafter, it could be determined how many moles of C, O, N, H and S from the feedstream would end up in which product. Lastly, averages were calculated when there was more than one element in a product. For example for CO the moles of C and O were summed up and divided by two. To calculate the vol% of CO this average was divided by the sum of all average molar quantities of all product gasses. For the higher hydrocarbons, not the ratio-method was used, but CH_4 relationships as given in Table 13 (van der Meijden, 2010).

For the contaminants H_2S, COS, NH_3 and HCl (HCN was not included in the model) the following assumptions based on experimental data were used from table 7.6 in van der Meijden, (2010):

- S: Of the total sulphur amount in the biomass, 60% is converted into H_2S and 6% to COS (no CS_2). From the remaining sulphur, 23% is part of the char, 6% is converted to thiophenes and 5% ends up in the ash.
- Cl: 20% of the chlorine in the biomass is converted into HCl. 60% goes to the char and the remaining 20% is part of the fly-ash.
- N: of the nitrogen present in the fuel, 50% is converted to NH_3, 30% to N_2, 10% to HCN and the remaining 10% resides in the char. The amount of N in the tar is neglected.

Tar production is taken from the ECN model as 20 g/Nm^3 of gas. It is unclear what an Nm^3 of gas means in the model, but as the input of air and CO_2 are set at 0°C a Nm^3 of gas is defined as the normal volume at 1 bar and 0°C (22.4 dm^3/mol).

3.4.3. Method validation

With the method as described above a re-calculation of the product gas composition in van der Meijdens' (2010) model was made to check if the method works appropriately. Therefore the ultimate analysis of the white labee pallets – input in van der Meijdens' model– was used as input for our gasification model so the gas composition after gasification was obtained. The results could be compared to the outcomes of the model of van der Meijden (2010) by comparing the outcomes with van der Meijdens' wet product gas composition.

3.4.4. Fuel cell

More specific information is needed about for instance the gas inlet requirements and operation parameters. Because SOFC fuel cells are used more often in combination with gasifiers (Bocci, Di Carlo, et al., 2014; Broust, 2008; Caliandro et al., 2014; Nagel, 2008), SOFC suppliers were contacted, see a list of suppliers and fuel cell characteristics in Table 24 of appendix C. First we contacted all suppliers by mail, followed by reminders after two weeks and after three weeks, plus a phone call when no response came. Because this method resulted in only one response that only could provide data of a fuel cell still in development phase (the Convion fuel cell), we eventually decided to base this study on a generic SOFC rather than on a specific brand.

The fuel cell operation parameters were based on a publication from the firm Strategic Analysis made for NETL (James et al., 2012). The NETL publication made a cost analysis of stationary fuel cell systems and specified quite well the parameters they used, based on data from the suppliers Fuel Cell Energy, Ceramic Fuel Cell Limited and NexTech. For the gas inlet requirements of the fuel cell, literature data (Aravind & de Jong, 2012; Bocci et al., 2014; van der Meijden, Veringa, Vreugdenhil, & Drift, 2009; van Paasen et al., 2006) were combined with information from Convion.

3.4.5. Cleaning and other equipment

Because both the gasifier and fuel cell work on high temperature, we first chose to design a hot gas clean-up system. In that way, the sensible heat of the gasses does not get lost, which increases the efficiency of the system (Aravind & de Jong, 2012; Bocci, Di Carlo, et al., 2014; Caliandro et al., 2014). So we chose cleaning options that could work on high temperature, plus other necessary equipment such as a biomass dryer and heat exchangers.

Then a meeting with Berend Vreugdenhil from ECN[7], who works on the MILENA gasifier, resulted in a more complete first system design. Changes in the order of equipment were made after follow-up email conversations, resulting in a second version of the system. Eventually, a meeting with Martin 't Hoff from Royal Dahlman further improved the system and the final system design was obtained.

3.5.   System analysis

This section focuses on the second research question that deals with the energy efficiency and electricity yield of the process. To calculate these outputs a model was created that could calculate the mass and energy flows of each part of the system. It was chosen to model the system completely in excel. The system design that is followed here is described in section 4.2.

3.5.1. Mass balances

Mass balances had to be created first because the mass flows (in kg/hr) were used as input for the energy balances later on. There was not one main method that could be adopted to calculate the mass balances, so the different ways that were used to obtain the mass balances are discussed here.

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[7] Summaries from interviews/meetings can be found in Appendix D

Dryer
For the dryer, the mass balance was based on the fact that the water content of the WTP had to be reduced from 60% to 25% moisture before entering the gasifier. The mass balance was solved with the total mass balance and partial mass balance for moisture in WTP.

Gasifier
The gasifier mass balance was constructed with as starting point the conversion ratio (kg/kg) of biomass to producer gas (ash free, with & without tar) derived from van der Meijden (2010). With this ratio, the gas flow of producer gas (ash free, with & without tar) in kg/hr was obtained. Tar mass flow was derived from the tar production in Nm^3/hr as given in the model plus the biomass to tar ratio. Ash mass flow was derived by combining the WTP input stream (kg/hr) and the ash content (wt%) from the ultimate analysis of WTP.

  The char mass flow was calculated with help of the ECN combustor datasheet (table 7.20 in the thesis of van der Meijden). The mass flow of ash and char were combined and the ash (wt%) content was given, so the separate ash and char flow could be calculated. Next, the ratio of char over the biomass inlet was calculated and used to determine the char mass flow from WTP gasification. Besides, 0.8% of the gas from the gasifier (so only raw producer gas) was recycled to the combustor as there is some 'leakage' between the two compartiments (van der Meijden, 2010). As a last note, CO_2 was used as fluidizer instead of O_2 to solve the CHO balance.

Combustor
To obtain the combustor mass balance, we assumed complete conversion of tar and char to flue gas. Ash flows from different input streams were combined in one output stream. One percent of the flue gas was send back to the gasifier, which is the typical leakage (van der Meijden, 2010). Lastly, 0.8% producer gas came as input to the combustor.

  To calculate the amount of air needed for full combustion it was assumed that all C in tar & char is converted to CO_2 and all H to H_2O. Then from the model we used that 99% of S in char is converted to SO_2 and 99% of Cl to HCl, 19% of N in char to NO and 2% of N in char to NO_2. The remainder of S, N and Cl is assumed to end up in the ash. Then the amount of moles of oxygen could be calculated via the molar ratio of C/N/S versus O in the combustion reactions. Afterwards, the amount of air needed was calculated according to the air composition (see Appendix E). Finally, the amount of air used for combustion was calculated with a lambda (stoichiometric air-to-fuel ratio) of 1.3 based on van der Meijden (2010).

Heat exchangers
In heat exchangers it was assumed that the mass flows of input and output streams are the same.

Cleaning equipment
In all types of cleaning equipment, some components in the gas were converted to other gasses, or completely removed and became part of a waste stream. To be able to calculate new product gas compositions, every component in the gas was first calculated from vol% to kg/hr. This was done by dividing vol% by the molar mass (kg/mol) of a specific component to obtain the kg/mol gas for each component. These were added up to a total of kg/mol for the total product gas. Then the ratio kg/mol for one compound was divided by the total kg/mol and multiplied with the mass stream of product gas (tar and ash free). The kg/hr mass flows of each component could then be modified based on the type of cleaning equipment. Conversion rates for each piece of equipment are given in Appendix E.

SOFC
For the fuel cell only an energy balance was calculated, the exact outlet mass flows were in this case not needed, as the electric efficiency was known.

3.5.2. Energy balances

First, energy balances were made for each piece of equipment separately.
To do so, the mass flows were converted to kg/s and energy streams were given in kW for both LHV and HHV. For each part of equipment, the same procedure was followed that only differed per type of energy stream:
- Energy in fuel/gas/products

The energy of the fuel (WTP) was calculated with help of the LHV and HHV of WTP with 60% and 25% moisture (see appendix E). The energy in gas was calculated by multiplying the mass flows of each component (kg/hr) with the energy content of each component (kJ/kg) for both HHV and LHV (see appendix E).
- Sensible heat

For sensible heat (Q_{sens}) calculations equation 6 was used.

$$Q_{sens} = c \cdot m \cdot \Delta T \qquad \text{(Eq 6)}$$

Where c is the specific heat capacity of a substance (kJ/kg/K), m the mass flow (kg/s) and ΔT the temperature difference (°C or K).
For air the specific heat capacity was derived from Cengel & Boles (2014). The specific heat capacity of the other gasses was derived with Aspen Plus. A heater block was fed with a specified gas composition (from the gasification model in excel) and heated up by one degree from different starting temperatures. In this way, the specific heat capacity at different temperatures was derived, as well as trend lines to predict the c at other temperatures.
For solid matter it was not possible to use this method. Therefore, the c for WTP, tar and char + ash were derived at one specific temperature from the model of van der Meijden (2010) and assumed to be constant with varying temperature. See appendix E for an overview of all c values used.
- Heat from evaporation of water (Q cond. Water)

The evaporation of water gives the amount of energy needed to evaporate the water that is present in certain products. It is thus the difference between the LHV and HHV of a fuel/gas/product stream.
- Heat supply/loss

When heat supply or loss is part of the energy balance, it is generally determined by difference to solve the energy balance. For the gasifier heat balance this was different, as the heat supply came from the combustor.

The exception to this procedure is the SOFC. For this apparatus, an electric and heat efficiency were assumed, based on Ammermann et al. (2015) & James et al. (2012). The electricity production was calculated by multiplying the product gas combustion enthalpy (in kW) with the electrical efficiency and/or efficiency to heat of the SOFC.

3.5.3. Heat integration

Through integration of heat flows the efficiency of the system will increase. Instead of obtaining heat for heating or cooling from outside sources, heat flows are as much as possible matched within the system. One method to do this is with an extensive pinch analysis (Nieuwlaar et al., 2014). Though, in this research we chose to perform a more basic matching of heat sources and sinks to get a first idea of the heat integration possibilities. To do so, the heat in or outflow from all heat exchangers and temperature rise and fall were combined in one table, including the energy content of streams.

This information was enough to see which heat streams could possibly be integrated, with as rule of thumb a minimum 20°C difference between the lowest temperature of the hot stream and the highest temperature of the cold stream. The heat that can be extracted by the fuel cell flue gas is calculated by multiplying the energy content of the completely cleaned product gas with the efficiency to heat of the SOFC given in Table 15. For any low quality heat that is left, it is assumed that this heat can be utilized in the district heating system of the AEB as long as the temperature is higher than 80°C.

3.5.4. Total system efficiency

Finally, we created a total energy balance by summing up all energy in- and outflows (LHV). From this balance, only the main in and outflows were used to calculate the electrical efficiency (Eq 7) defined as the electricity output (in kW) divided by the sum of all energy inputs (Fuel, heat, electricity). Equation 8 includes electricity and the useful heat as output to obtain the total energy efficiency.

$$\eta_{elec} = \frac{E_{elec,out} - E_{elec,in}}{\sum E_{fuel} + E_{heat,in}} \qquad\qquad \text{(Eq 7)}$$

$$\eta_{total} = \frac{E_{elec,out} + E_{heat,out} - E_{elec,in}}{\sum E_{fuel} + E_{heat,in}} \qquad\qquad \text{(Eq 8)}$$

3.6.   Economic analysis

The economic analysis has two important objectives. The first one is to determine if a certain project is economically feasible. The second one is to be able to compare the outcomes with other possibilities, such as fermentation of WTP instead of gasification. Economic indicators that are typically used for this purposes are the discounted Pay Back Period (PBP), Net Present Value (NPV), Internal rate of return (IRR) and the Levelized Costs of Electricity (LCOE), see Table 3 for the formulas (Blok, 2007; Short, Packey, & Holt, 1995). These indicators were chosen because they all give a slightly different insight.

The PBP is an easy to interpret number for an investor, though it is not meant to for comparing projects with for example a different size (Short et al., 1995). The other parameters can used to compare alternatives. The NPV shows the value of the project over the project lifetime, so the profit earned or investment lost. The IRR gives the rate of return one can expect on the investment. The most relevant for this project is the LCOE, as it clearly shows the costs of producing electricity with this technology and can easily be compared to other technology options. The values obtained as input data for the different economic indicators are given in Table 4. How they were determined is explained in the next paragraphs.

Table 3 Formulas and short explanation of economic indicators.

Indicator	Explanation	General Formula	Symbol explanation
PBP	Indicates from which moment on the generated benefits of a project can be seen as profits.	$I = \sum_{i=1}^{PBP} \dfrac{B - OM - F}{(1+r)^t}$	I = Total Plant Cost (€) B = Annual benefits (€/y) OM = OPEX - operation and maintance cost (€/y) F = fuel cost (€/y) r = discount rate (%) t = time (y) L = lifetime of the project (y) E = electricity production (kWh/y)
NPV	Present value of a project taking into account all current and future costs and benefits. NPV > 0 means the project will generate profit	$NPV = -I + \sum_{i=1}^{L} \dfrac{B - OM - F}{(1+r)^t}$	
IRR	The rate of return you can expect within the project lifetime (discount rate when NPV = 0)	$I = \sum_{i=1}^{L} \dfrac{(B - OM - F)}{(1+IRR)^t}$	
LCOE	Cost of electricity over the projects lifetime	$LCOE = \dfrac{I + \sum_{i=1}^{L} \frac{OM+F}{(1+r)^t}}{\sum_{i=1}^{L} \frac{E}{(1+r)^t}}$	

Table 4 Economic input data

Economic Data	
Total Plant Costs = I	€ 32,673,043
OPEX/OM costs = OM	€ 911,457
Fuel costs (WTP) = F	0 €/tonne
Annual benefits = B	€ 1,063,795
Electricity production (GWh) = E	20.21 GWh
Project lifetime = L	20 years
Discount rate = r	5%
Availability	80%
Electricity price (production)	49.17 €/MWh
Heat price (production)	5 €/GJ
Exchange rate (dollar/euro)	0.9142 €/$
Plant construction time	1 year
Subsidy	0.1038 €/kWh

3.6.1. Total Plant Costs

In this research it was chosen to calculate the Total Plant Costs (TPC) that include the equipment costs, Engineering, Procurement & Construction (EPC) and contingencies (Rubin et al., 2013). The ECP costs include labour during construction, civil work, building, site management etc. Contingencies are added for unforeseen cost in the project development or during building of the plant.

The TPC does not include owners cost such as costs for land, insurance and taxes, working capital and start-up of the plant, which would give the Total Overnight Costs. Also the time that is needed for building the plant after investments are done is not taken into account. These costs are defined by Rubin et al. (2013) as interest during construction and cost escalation during construction. Including these factors would have result in an estimate

for the Total Capital Requirement. We decided that for a feasibility study such as this one, the Total Plant Costs include sufficient information. For the gasifier system and cleaning equipment, direct cost data were obtained from Royal Dahlman. This data (Table 5) include all factors for total plant cost at the required size of 6MWth[8] input for the gasifier.

Table 5 Build-up of CAPEX (capital expense)/Total Plant Cost system for the dryer, gasifier and cleaning system, obtained from Royal Dahlman and based on the 4MW biomass to SNG plant that will be built in Alkmaar.

Scope & Budget Estimate MILENA (6MW) - OLGA - AQUA - HDS -ZNO- steam injection-optional Pre-reformer	CAPEX
MILENA-OLGA	
Biomass storage & drying	€ 720,000
Total MILENA-OLGA & water scrubber	€ 8,350,000
Cleaned product gas compressor	€ 520,000
HEX HDS HEX ZnO HEX Steam mixer HEX	€ 1,900,000
optioneel Pre-reformer plus HEX	€ 240,000
Balance of plant (BOP) scope & installation work	€ 2,500,000
Civil works, building	€ 700,000
EPC/project & site management	€ 1,500,000
Total	***€ 16,430,000***
Contingency; 10%	*€ 1,643,000*
TPC (without SOFC)	**€ 18,073,000**

Though, the TPC are not complete yet without the SOFC investment costs. No specific cost data could be obtained from suppliers directly and therefore the following method was used based on Aranda, Drift, & Smit (2014) & Sinnott (2005).

1. Equipment cost estimation
   SOFC costs were retrieved from as many public sources as possible (Ammermann et al., 2015; Battelle, 2014; James et al., 2012; Mcphail, Leto, & Boigues-Muñoz, 2013; Schoots, Kramer, & Zwaan, 2010; Thijssen, 2007; Weimar, Gotthold, Chick, & Whyatt, 2013).
2. Scaling of the equipment to the required size
   For the scale of the system, see 4.3.3. For an SOFC no scaling factors were found, but based on some general information on scaling from (Towler & Sinnott, 2013) it was decided to choose a scaling factor of 0.85. Because fuel cell systems are made by assembling many small cells to into stacks to obtain a large system, the costs will not be reduced that much when a larger system is made. Only the stack packaging will be strongly reduced when producing larger systems (Thijssen, 2007).
   Formula used for scaling:

$$C_2 = C_1 \cdot \left(\frac{S_2}{S_1}\right)^n \qquad\qquad \text{(Eq 9)}$$

   Where C_2 are the scaled investment costs with capacity S_2 and C_1 the investment costs with capacity S_1. n is the scaling factor (0.85 in this case).
3. Converting to euro
   When prices are given in US dollar, they were converted to euro-2015 by using conversions factor from OANDA (2016). Specific conversion rates used can be found in Appendix G.

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[8] After drying the energy content of the WTP is increased (see Table 18), hence the 6MWth instead of 5MWth scale.

4. Converting the investment costs to 2015
When costs are derived from publications, cost data can be old and therefore it is needed to correct for inflation. Typically the Chemical Engineering Plant Cost Index (CEPCI) values are used for engineering purposes. Though, those values are based on US data. Therefore, in this research we chose to use the Consumer Price Index (CPI) from CBS (CBS, 2016).
Formula used for conversion:

$$C_2 = C_1 \cdot \left(\frac{CPI_2}{CPI_1} \right) \qquad \text{(Eq 10)}$$

5. Location factor
For prices that were at first instance given in dollar, a location factor of 1.23 was used. This value is based on the value of 1.19 given in (Towler & Sinnott, 2013) for US Gulf Coast to the Netherlands in 2003 and updated with dollar to euro conversion rates in 2003 and 2015 (see appendix G).
6. Cost escalation
For fuel cells one factor that include all EPC costs was not found. Instead, an installation factor could be derived of 1.42, which is typical for SOFC systems (James et al., 2012; Thijssen, 2009; Weimar et al., 2013). According to Ceasar (2011), equipment cost and installation sum up to the Total Direct Plant Cost. To arrive at the EPC costs an 'indirect cost' factor of 14% (so 1.14) had to be added that included yard improvement, service facilities, engineering/consultancy cost, building and a miscellaneous factor (Ceasar, 2011). So cost escalation with these two factors was done to arrive at the EPC costs of the SOFC.

In appendix G an overview of the cost data about SOFCs is given. Finally investment including installation costs were set at € 11,853,304 and indirect costs of €1,659,463 were added. To obtain the final TPC the ECP for of the gasifier & cleaning system and ECP for the SOFC were combined. Then a 10% contingency factor was added to make up for all unforeseen costs, based on the cost specifications obtained from Royal Dahlman. Results are shown in Table 6.

Table 6 Total Plant Cost complete system

Type of equipment	CAPEX
Biomass storage & drying	€ 720,000
Total MILENA-OLGA & water scrubber	€ 8,350,000
Cleaned product gas compressor	€ 520,000
HEX HDS HEX ZnO HEX Steam mixer HEX	€ 1,900,000
Balance of plant (BOP) scope & installation work	€ 2,500,000
Civil works, building	€ 700,000
EPC/project & site management	€ 1,500,000
Fuel Cell + installation	€ 11,853,304
Fuel cell indirect cost	€ 1,659,463
Total	**€ 29,702,767**
Contingency; 10%	*€ 2,970,277*
Total Plant	**€ 32,673,043**

3.6.2. Annual benefits

Benefits are mainly earned by selling electricity; also some heat can be sold. The revenues for heat were obtained via the AEB and set at 5€/GJ of heat. The revenues for selling electricity in the Netherlands were determined by taking the average over the last 10 years (2005-2015) of the index that can be downloaded from the APX website[9] (APX, 2016) and multiplying with the electricity production. This average was 49.17 €/MWh or 4.917 eurocent/kWh. This selling price was held constant during the total analysis in this study. The annual benefits were only used for calculation of the PBP, IRR and NPV. For the LCOE the annual benefits were not relevant as the 'required benefits' are calculated there, see Table 3.

3.6.3. Operating cost (OPEX)

The operation costs or operation expense (OPEX) include both fuel and operation & maintenance (O&M) costs of the system. The O&M costs can be divided into fixed O&M costs such as operating labor, supervision, maintenance, rent of land etc. (Towler & Sinnott, 2013). The variable O&M costs are the costs that are proportional to the plant output or operation rate and include for example raw materials, consumables, utilities (air/steam/cooling water etc) and effluent disposal (Towler & Sinnott, 2013).

For this report, the OPEX for the gasifier and cleaning system was derived from Royal Dahlman as 81.40€/h or €713.064/year. The following components are included:
-   Labour in 3-shifts of 8h each, 3 people during the day (2 shifts) and 2 at night (1 shift). Including weekends and holidays this will be 13 fulltime-equivalents (FTE) in total per year.
-   Oil consumption OLGA (although this negligible)
-   Bed material
-   Electricity consumption
-   Natural gas
-   Waste water treatment
-   Catalyst
-   Absorbents
-   General plant maintenance
-   Ash disposal
-   Other not reported consumables (such as demi water for steam, disposal / regeneration of ZnO, lubricants etc.)

For the fuel cell, an estimation of the OPEX was taken from Ammermann et al. (2015) as € 198,381.72/y. Furthermore, the generic SOFC used in this project has an expected lifetime of 16 years with 3 stack replacements in between. It was assumed that new stack replacements have to be done after 4 years, 8 years and 12 years. Stack replacement costs are € 3,860,586 (Ammermann et al., 2015).

Normally the fuel costs form a large share of the OPEX, but in this system the fuel costs (WTP) are set at 0 €/tonne. At the moment, Waternet has to pay 70 €/tonne for WTP removal, so actually the fuel costs are even negative. Though, when a profitable business case is made out of WTP, Waternet will not be inclined to pay for the waste toilet paper they deliver. Therefore the fuel costs were set at 0 €/tonne.

3.6.4. Other parameters

Another important parameter is the discount rate, which was obtained from the AEB as 5%. Because the AEB will build the system, it was the most logical to ask them about the discount rate they use. Lastly there is the project lifetime, which was in this case based on the lowest expected lifetime of an important part of equipment; the fuel cell. The generic fuel cell chosen for this project has a expected lifetime of 16 years (Ammermann et al., 2015). The other important part of equipment, the gasifier, can work for 30 years, but for the project lifetime of gasifier systems it is more typical to take 10-15 years (Asadullah, 2014; Boerrigter et al.,

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[9] If you follow the link in the reference, the file can be downloaded via a column on the right side of the webpage

2005; Zwart et al., 2006). Based on this knowledge we decided to restrict the project lifetime to the fuel cell lifetime, so 16 years.

3.6.5. Subsidy

The costs for the system are expected to be relatively high as both the gasifier technology and SOFC are still not fully commercialized yet. Therefore, it is likely that government subsidy will be required to make the project profitable. In the Netherlands there is a subsidy scheme called Stimulation for Renewable Energy Production (SDE+) that might be applicable for this project (RVO, 2016). In 2016, eight billion euro is available for renewable energy projects (Kamp, 2015). Subsidy is given per kWh of energy produced and should be just high enough to overcome the economically unfeasible part of electricity production from renewable sources.

The subsidy values are divided over different categories, both in terms of the type of energy (electricity, gas, heat or combined heat and power) as well as type of generation (hydropower, photovoltaic, wind energy, geothermal energy, energy from waste water treatment, burning or gasification of biomass, fermentation of biomass) (Lensink & van Zuijlen, 2014). This project falls in between categories. Of course electricity is the main goal, but some heat is generated as well and it is also possible to produce gas. Regarding the type of generation, it could be energy from waste water treatment as well as gasification of biomass.

A summary of the most plausible options and subsidy values is given in Table 7, and were verified with Sander Lensink, SDE+ expert from ECN. Another possibility would be that the combination of biomass gasification with a fuel cell is accepted as a new or 'free' category. Then the subsidy value can be different. For now the maximum value (0.1038 €/kWh) was chosen, to know if this amount would be enough to make the project economically feasible.

Table 7 Possible subsidy values

Renewable energy technology	Type of energy	Subsidy value (€/kWh)
Biomass gasification	Gas	0.106 – 0.022= 0.1038
Thermic conversion <100 MWe	CHP	0.077 – 0.023 = 0.054
WWTP sludge fermentation	CHP	0.060 - 0.032 = 0.028
WWTP gas production from sludge	Gas	0.032- 0.022 = 0.010

Source: Letter of the minister of Economic Affairs (Kamp, 2015), who based his numbers on the ECN advice (Lensink & van Zuijlen, 2014). The values are given as basis-subsidy values and an expected correction value, subtracting the latter from the former gives the expected subsidy value. The ECN advice gives more information about this.

3.7. Learning curves

The general definition of learning is that the costs of producing something will decrease with a constant factor by each doubling of the production (Blok, 2007). There can be different reasons for a cost decrease and thus different types of learning exist. Pure learning means that the knowledge and working experience of the production of a technology expand when the production is increased (Rivera-Tinoco, Schoots, & Zwaan, 2012). Pure learning is applicable to building as well as operation of a plant (Aranda et al., 2014). Another factor can be economies of scale, which means the production costs decrease when bigger units are produced, as was discussed earlier (in section 3.6.1). Two helpful formulas when working with learning curves are:

$$C_t = C_o \cdot \left(\frac{P_t}{P_0}\right)^{-\alpha} \qquad \text{(Eq 11)}$$

$$lr = 1 - 2^{-\alpha} \qquad \text{(Eq 12)}$$

Where C_t is the cost at a produced or installed capacity P_t of a technology, And C_0 the current produced or installed capacity P_0 of a technology. α is the learning index, $2^{-\alpha}$ is the progress ratio and lr is the learning rate that is usually expressed as a percentage.

To illustrate the concept, a learning rate of 10% means that after each doubling of production the price of the technology decreases with 10%. The progress ratio will then be 0.9 and the learning index will be 0.152. For most technologies, learning rates are between 5% and 30% (Blok, 2007).

In this research, the learning curve analysis was done for the SOFC and the gasifier system (which included MILENA-OLGA & water scrubber). First, the total investments on the gasifier system and the fuel cell system (including stack replacements) were divided by the total electricity production to determine their share in the LCOE. Then their specific share was varied by applying the learning effects, while keeping the remainder of the LCOE constant. In order to make the analysis, a proper learning rate (lr) had to be determined, as well as the current cumulative installed capacity of the technology (P_0), which is done in the next section.

3.7.1. Fuel cell

In order to find out which learning rate to choose for the SOFC, an overview of learning rates was created based on literature (Iyengar, Keairns, Krulla, & Newby, 2013; Rivera-Tinoco et al., 2012; Schoots et al., 2010; Thijssen, 2009; Weimar et al., 2013) and can be found in Appendix H. The values were compared based on the type of learning they include in the learning rate and to what extent the learning rate was applicable on a large stationary SOFC system, the learning rate chosen was 15%.

The cumulative installed capacity of SOFC was derived by combining data from a sample report from Grand View Research (2014) with global SOFC shipment data from Carter & Wing (2013) and Hart, Lehner, Rose, Lewis, & Klippenstein (2015). Finally, cumulative global shipments of 194 MW were used in the calculations.

3.7.2. Gasifier system

The learning rate of biomass gasifiers and installed capacity was obtained from three different sources (Aranda et al., 2014; van den Broek et al., 2009; Knoope, Meerman, Ramírez, & Faaij, 2013). The data were assessed based on the applicability on a biomass gasifier system. Finally a 10% learning rate was chosen and a global installed capacity of 20GW, for more information please see appendix H.

3.8.   Uncertainty analysis

An uncertainty analysis of the results of a research can be very useful to investigate how robust your results are. It can also provide you with information about which aspects of your research have priority when you want to come up with solid conclusions. As this study is a first feasibility study on a WTP to electricity plant, it is of utmost importance to 'know what you do not know'. If you become aware of the less solid parts of your analysis, you can also start improving them.

In this study, two methods for uncertainty analysis were combined and applied on the model input parameters. First, a sensitivity analysis was done to find out to what extend changes in a certain input parameter would influence the LCOE. Subsequently, a pedigree-analysis was done to determine how strong all input parameters are, which means to assess how sure we are about them. Then finally a diagnostic diagram was created that combined information from both types of analysis and draw conclusions about the priorities for further research.

3.8.1. Sensitivity analysis

In the sensitivity analysis, selected input parameters were varied over a certain range to assess the influence they had on the LCOE. The parameters were chosen based on their expected influence on the results. The range over which a parameter was varied highly depended on the way the parameter was obtained. How the ranges were chosen is explained below Table 8.

The analysis itself was done by using the Data Table function in excel for most parameters, but this was not possible for the WTP composition, learning rate of the SOFC and lifetime of the project. For the WTP composition, two slightly different WTP compositions were

used as model input, sample 2 and 3 from Table 10. Large differences in input were not possible because then the CHO balance is not solved.

The sensitivity of the SOFC learning rate was determined by plotting different learning rates in one graph. Then values were obtained from the point were the learning reaches a plateau (in this case 100GW), then the variation in values was assessed at this point. For the project lifetime, the (discounted) cash flows of the system were adjusted to lifetimes of 10, 15, 25 and 30 years and then the LCOE was calculated again. For a 10 year system lifetime only two stack replacements were done, it was not needed to invest in a new fuel cell. For a lifetime of 20 years, an new SOFC was bought after 16 years, and sold at half the price four years later when the project was finished to earn back part of the capital. For the 25 and 30 year project lifetimes, a second SOFC was bought after 16 years, but no stack replacements or investments were done afterwards.

Table 8 Parameters for sensitivity analysis plus the range over which they will be varied.

Parameter	Current value	Range
Amount of WTP (ton/y)	25000	20000-30000[a]
Water content WTP (%)	60	50-70[b]
WTP composition		2 different compositions based on different values of the ultimate analysis
Price of WTP (€/tonne)	0	-70 to + 40[c]
SOFC investment costs (M€)	11.85	2.5-17[d]
Efficiency SOFC (%, LHV)	55	45-70[e]
Scale factor SOFC	0.85	0.8-0.95[f]
Learning rate SOFC (%)	15	10-25[g]
Lifetime of project (y)	16	10-30[h]
Discount rate (%)	5	3 -10[i]
OM costs Gasifier (€/y)	713,064	550,000 – 1,100,000[j]
OM costs – SOFC (€/y)	198,394	160,000 – 300,000[j]
value (eurocent/kWh)	10.38	1.0 – 15[k]

[a] +/- 20% the production rate (Towler & Sinnott, 2013)
[b] Variation found in the STOWA/Waternet report on WTP (Ruiken et al., 2010)
[c] – 70 is the current price Waternet has to pay to process the WTP, 40 euros is the price of one tonne wood chips in Europe when energy content is adapted to WTP (EUBIONET3, 2011; Hoefnagels, Junginger, & Resch, 2015; Prislan, Krajnc, Jemec, & Piskur, 2014)
[d] Lower and higher range in from Table 50.
[e] Lower range (Leo, n.d.; Mcphail et al., 2013), higher range (Mcphail et al., 2012).
[f] These values are educated guesses, based on the fact that stacking separate small fuel cells to create a big system will not result in high scale factors.
[g] Based on the high and low ranges in Table 51.
[h] Lower range: based on lowest lifetime of similar projects (Asadullah, 2014; Boerrigter et al., 2005; Zwart et al., 2006), higher range is expected lifetime of the gasifier (Royal Dahlman, personal communication).
[i] Lower range is discount rate for government projects, higher range for commercial projects (Short et al., 1995). The AEB is a government owned company and could therefore fall within this range.
[j] lower range -20%, higher range plus 50% (Towler & Sinnott, 2013)
[k] Lowest value from Table 7, 15 cents the upper limit according to Kamp (2015).

3.8.2. Pedigree analysis

To get more information about the strength of the input parameters, we performed a pedigree analysis. This analysis assesses the strength of the input parameters, helping prioritize the next steps to a more detailed feasibility study. Strength in this case means certainty on a given parameter value.

In this analysis, different experts assign scores to each input parameter, based on a number of so called 'pedigree criteria'. The tool used to score each parameter is a pedigree

matrix, where for each parameter and criteria a score from 4 (highest) to 0 (lowest) can be given. The practice of pedigree analysis originates from environmental modelling (Refsgaard, van der Sluijs, Højberg, & Vanrolleghem, 2007; Sluijs, Risbey, & Ravetz, 2005; Sluijs, Craye, et al., 2005) but can also be used as part of the uncertainty analysis of process models for (chemical) engineering, as was recently shown by van der Spek, Ramirez, & Faaij (2016). In this study, most data are derived from either experts at institutes/suppliers or from literature. Therefore, the pedigree matrix (Table 9) is designed to assess the strength of parameters from this kind of source.

Table 9 Pedigree matrix for the assessment of model input parameters and model strength. The matrix is based on unpublished work of van der Spek (2016), who based his table on van der Sluijs et al. (2005) and (Weidema, 1998).

Strength	Proxy	Reliability of source	Completeness	Validation process
4	A direct measure of the desired quantity	Measured/official industrial, vendor, and/or supplier data	Complete data from a large number of samples over a representative period	Compared with independent data from similar systems that are already built or with independent measurements of the same variable over long domain
3	Good fit to measure	Qualified estimate by industrial expert supported by industry data	Complete data from a large number of samples but for unrepresentative periods or from representative periods but for a small number of samples	Compared with independent data of similar systems that have not been built or with independent measurements of closely related variables over shorter period
2	Correlated but does not measure the same thing	Peer reviewed data derived from independent open literature	Almost complete data from a smaller number of samples or for unrepresentative periods or incomplete data from adequate number of samples and periods	Validation measurements are not independent, include proxy variables or have limited domain
1	Weak correlation but commonalities in measure	Non-reviewed data derived from open literature	Almost complete data but from a small number of samples and unrepresentative periods	Weak and very indirect validation
0	Not correlated and not clearly related	non-qualified estimate or unknown origin	Incomplete data from a small number of samples for an unrepresentative period	No validation performed

The pedigree-matrix was used during a workshop with three PhD-researchers from the Energy & Resources group from the Copernicus Institute (Utrecht University) who all had experience with pedigree analysis. Also, five experts from Waternet, the AEB, Royal Dahlman and ECN were invited because they had specific knowledge about (a part of) the project. In the end, two experts (from Waternet and Royal Dahlman) did attend in the pedigree analysis workshop. The workshop consisted of a project presentation, discussing the pedigree analysis method and individually scoring all parameters with the help of an information document.

3.8.3. Diagnostic diagram

A diagnostic diagram combines the input from the sensitivity analysis and the pedigree analysis in one graph. The x-axes shows the parameter strength, while the y-axes gives the parameter sensitivity. To create the diagnostic diagram, the sensitivity results were first normalized to the highest sensitivity score. This was done for each parameter separately by determining the maximum deviation from the LCOE to both sides of the graph -decreasing and increasing LCOE - and dividing by two. Then the highest average deviation from the LCOE was set at one and all other parameters were normalized to this number and used as values on the y-axes. For the pedigree score, the median of all input values for a certain criterion within one parameter was calculated. Then, the average was taken over all criteria for one parameter to obtain one final pedigree-score for each parameter that could be used as value on the x-axes.

4. Results

4.1. Fuel Characterization

Figure 11 shows the WTP before and after pulverization in a blender, to provide an idea of the appearance of the fuel.

Figure 11 Left: WTP derived from the sludge at WWTP Horstermeer washed and dried. Right: WTP after pulverization in the blender.

4.1.1. Ultimate analysis

The ultimate analysis was done of three WTP samples at *Mikroanalytisches Labor Kolbe* (Germany). Results of the three samples and standard deviation are given in Table 10, the oxygen content was calculated by difference.

Table 10 Results of the Ultimate analysis of WTP. Oxygen is calculated by difference.

	Sample 1	Sample 2	Sample 3	Average	stdev
C (wt%)	42.2	43.74	42.41	42.78	0.84
H (wt%)	5.87	6.63	6.15	6.22	0.38
O (wt%)	46.5	45.45	44.23	45.39	1.14
N (wt%)	0.42	0.86	0.38	0.55	0.27
S (wt%)	0.03	0.02	0.86	0.30	0.48
ash (wt%)	4.98	3.3	5.97	4.75	1.35

4.1.2. Energy content

Table 11 shows the results of the WTP heating value calculation. The values in the table are quite similar, but the formula-results give a lower energy content of WTP than the ECN analysis. One the one hand, the formulas can only give an approximation, but on the other hand the WTP used for ultimate analysis that was used as input for these formulas is more representative than the material from ECN analysis. The reason is that the ECN analysis in the report of Ruiken et al. (2010) is of WTP that is extracted by means of sieving the water before it goes to the sedimentation tank, while the material in this study is extracted from the sludge after the sedimentation tank. For the current analysis, it was chosen to use the results of the Channiwala & Parikh formula.

Table 11 Energy Content (LHV and HHV) of WTP (dried, about 3.4 % moisture) obtained in different ways

Source	LHV (MJ/kg)	HHV (MJ/kg)
ECN analysis from Ruiken et al. (2010)	16.5	17.85
Formula from Milne et al. (1990)	15.93	17.29
Formula from Channiwala & Parikh (2002)	16.13	17.49

4.2. System design

4.2.1. Gasifier

After analysing the different gasifier options, it was decided that the indirect fluidized bed suits this project best. Based on the requirements in 3.4.1 we decided that in this project a medium temperature gasifier is preferred over a high temperature entrained flow gasifier. Entrained flow gasifiers work at large scales (5-100 MWe) and have a very limited fuel flexibility (Ruiz et al., 2013). In the medium temperature gasifier range, both the updraft fixed bed and fluidized bed gasifier fulfil the appropriate scale range, but the fluidized bed gasifier gives more fuel flexibility and somewhat lower tar production (Ruiz et al., 2013).

Besides, the indirect fluidized bed gasifier is the only option that can fulfil the last requirement (low N_2 content) without the need for an expensive air separation unit (Dascomb & Krothapalli, 2014; Mcphail et al., 2012; van der Christiaan M. van der Meijden et al., 2010; Nagel, 2008; van Paasen et al., 2006; Pfeifer et al., 2011). Therefore, an indirect fluidized bed gasifier was chosen as the best option in the WTP to electricity system.

There are three producers for indirect fluidized bed gasifiers: the Fast Internally Circulated Fluidized Bed (FICFB) gasifier, from Güssing Renewable Energy (Hofbauer et al., 1997), the MILENA gasfier from ECN/Royal Dahlman (van der Meijden, 2010) and the SilvaGas gasifier from Rentech (Paisley & Overend, 2002), see Figure 12 for illustrations.

The MILENA gasifier was selected for this project because it has a higher efficiency to synthetic natural gas than the FICFB (van der Drift, Zwart, Vreugdenhil, & Bleijendaal, 2010) due to a low steam to biomass ratio (0.1 for MILENA vs 0.7-1.1 for FICFB) (van der Meijden, 2010). Silvagas and MILENA are quite similar in gas composition and cold gas efficiency under equal conditions (van der Meijden, 2010) but the Silvagas gasifier is not well documented and was therefore not chosen in this project. Operation parameters for the MILENA gasifier are given in Table 12.
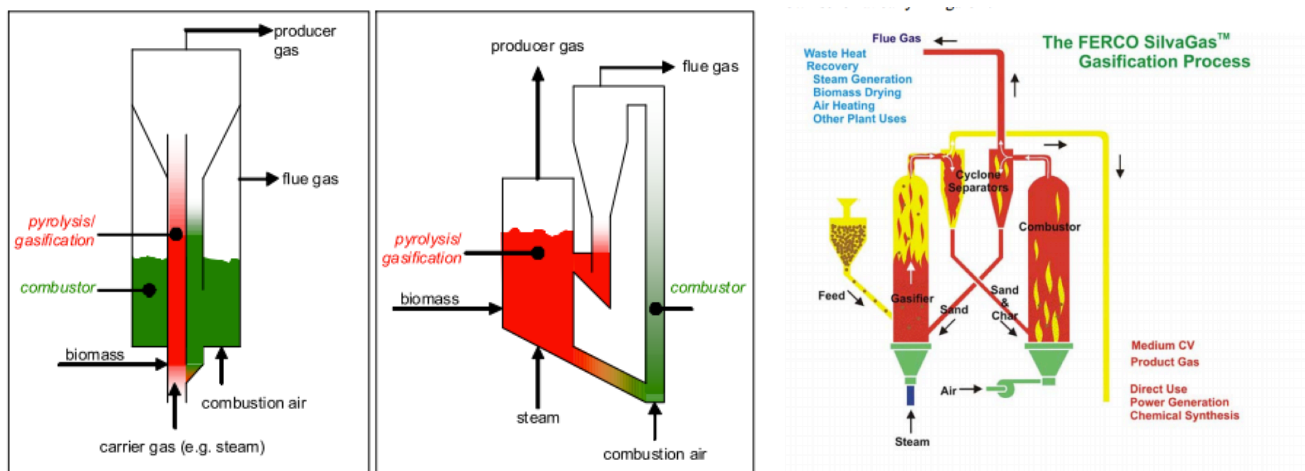
Figure 12 Comparison of the Milena (left), FICBF (middle) and Silvagas (right) process. Sources picture on the left (Van der Drift et al., 2010) , right (Paisley & Overend, 2002).

Table 12 Operation parameters MILENA gasifier

Parameter	Value
Operating temperature gasifier	850 °C
Outlet temperature producer gas (after settling chamber)	800 °C
Operating temperature combustor	900 °C
Operating pressure	atmospheric
Bed material	Sand

4.2.2. Method validation

The method to obtain the gasifier product gas composition was validated by checking if the method resembles the product gas composition from the model of van der Meijden (2010). Results are shown in Table 13. The method is clearly not perfect, but for the purpose of this research it is considered as good enough.

Table 13 Gas composition method validation. The third column gives the actual gas composition (wet) from the ECN model, while the re-calculation gives the results of the model that was made in this research. The last column shows the deviation (in percentage) between those two.

Output		Wet gas compostion - model van der Meijden (2010)	Re-calculation of the gas composition with own method	Difference	Relationships
CO	vol%	17.1%	16.7%	2.49%	
H2	vol%	18.3%	18.8%	-2.99%	
CO2	vol%	13.9%	13.5%	3.17%	
O2	vol%	0	0.0%	0.0%	
H2O	vol%	37.3%	37.5%	-0.49%	
CH4	vol%	8.1%	8.3%	-2.30%	
N2	vol%	1.2%	1.2%	-3.80%	
Ar	vol%	0.01%	0.0%	0.0%	
C2H2	vol%	0.2%	0.2%	-3.58%	CH4 * 0.025
C2H4	vol%	2.7%	2.7%	-1.28%	CH4 * 0.33
C2H6	vol%	0.2%	0.2%	8.85%	CH4 * 0.022
C3H6	vol%	0.06%	0.1%	-10.48%	CH4 * 0.008
C6H6	vol%	0.7%	0.7%	1.75%	CH4* 0.083
C7H8	vol%	0	0.0%	0.0%	CH4 * (0.093-0.0001*T)
H2S from fuel	ppmv	27	29	-8.60%	60% S in fuel
COS	ppmv	3	3	2.26%	6% of S in fuel
NH3 from fuel	ppmv	718	727	-1.27%	50 % of N in fuel
HCl from fuel	ppmv	11	9	19.62%	20% of Cl in fuel

4.2.3. Gasifier product gas composition

Using the method as described in 3.4.2 the gas composition for WTP gasification in the MILENA gasifier is given in Table 14. To solve the C,H,O balance, only CO_2 was added as no extra steam was needed.

Table 14 Raw product gas composition after MILENA for gasification of WTP

Raw product gas		Prediction	
CO	vol %	17.5%	
H2	vol %	17.1%	
CO2	vol %	14.3%	
O2	vol %		
H2O	vol %	37.1%	
CH4	vol %	7.9%	
N2	vol %	1.6%	30% of N in fuel + 1% recycling from combustor
Ar	vol %		
C2H2	vol %	0.20%	CH4 * 0.025
C2H4	vol %	2.6%	CH4 * 0.33
C2H6	vol %	0.17%	CH4 * 0.022
C3H6	vol %	0.06%	CH4 * 0.008
C6H6	vol %	0.66%	CH4* 0.083
C7H8	vol %	0.06%	CH4 * (0.093-0.0001*T)
H2S from fuel	Vppm	915	60% S in fuel
COS	Vppm	91	6% of S in fuel
NH3 from fuel	Vppm	3171	50 % of N in fuel
HCl from fuel	Vppm	810	20% of Cl in fuel
Thiophene	Vppm	91.5	6% of S in fuel
HCN	Vppm	637	10% of N in fuel
Tar	mg/nm3 wet	20000	

4.2.4. Fuel cell

The fuel cell chosen for this project is an SOFC, because this type of fuel cell can be combined with a gasification system (Bocci, Di Carlo, et al., 2014; Broust, 2008; Nagel, 2008) and has a relatively good resistance to contaminants in the product gas compared to other fuel cell types (James et al., 2012; Mcphail et al., 2012). A generic SOFC is used in this project that has operation parameters as given in Table 15 and with inlet requirements that can be found in Table 16.

Table 15 Fuel cell operation parameters.

Parameter	Value
Operating temperature	819 °C
Power density	291 mW/cm^2
Cell Voltage	0.8 V/cell
Operating pressure	~1.4 atm
Fuel Utilization	80%
Electrical system efficiency (HHV)	49%
Electrical system efficiency (LHV)	55%

Source: (James et al., 2012)

Table 16 Inlet requirements fuel cell used in this research

Contaminant	Max. allowable concentration
Particulates/dust	10-100 ppmv
Tar	< 2000ppmv
HCl	<5 ppmv (reported as 'few')
NH_3	Fuel for SOFC
HCN	Unclear
H_2S+COS+thiophene	< 1ppmv (or less, when possible)

Sources: (Aravind & de Jong, 2012; Bocci et al., 2014; van der Meijden, Veringa, Vreugdenhil, & Drift, 2009; van Paasen et al., 2006)

4.2.5. Total system design

Based on the discussion of equipment options in chapter 2, a total system design was created. The final design is shown in Figure 15 and resembles to a large extent the set-up for producing Bio-SNG from ECN with MILENA, OLGA and a large part of the ESME (ECN System for MEthanization) cleaning system (Van der Meijden, 2010; Rabou & Almansa Aranda, 2015). Though before arriving at this design, many other options have crossed the stage and therefore, not only reasons for the final choice Figure 15 will be discussed, but also two alternative earlier designs (Figure 13 & Figure 14).

Drying
First, the WTP has to be dried, because it contains about 60% moisture after extraction from the WWTP. For drying, a simple single pass rotary air drum dryer was selected (van der Meijden, 2010) that works on hot air for drying the WTP. An air temperature of 120°C should be sufficient, according to Martin van 't Hoff from Royal Dahlman (personal communication).

Gasifier
The gasifier of choice is the MILENA gasifier from ECN, as argued in 3.4.1. This system is well explained in the thesis of van der Meijden (2010) and also includes a heat exchanger and filter for flue gas from the combustor.

Particle removal
It was decided to use a cyclone for particle removal based on the system from van der Meijden (2010). The flue gas from the combustor is also filtered to remove the ash.

Tar removal
For tar removal, the OLGA system is chosen as it is well tested in combination with MILENA and the tars can be recycled and burned in the combustor, leaving no other waste in terms of catalysts or separate tar burning. Because the OLGA system inlet temperature is set at 400°C, a heat exchanger is required to cool the gas from the gasifier. In Figure 13 OLGA is placed at the end of the chain, so the gas only had to be heated up once after OLGA to fuel cell temperature. Though, email conversations with Berend Vreugdenhil (ECN, personal communication) revealed that OLGA should be placed directly behind the cyclone because the risk of tar condensation is too high. This suggestion was incorporated in the final design.

Nitrogen and sulphur cleaning
At first, it seemed most logical to use a hot gas clean-up system because the sensible heat of the gasses does not get lost, which increases the efficiency of the system (Aravind & de Jong, 2012; Bocci, Di Carlo, et al., 2014; Caliandro et al., 2014). Therefore, the first two line ups (Figure 13 & Figure 14) consisted of an $NaHCO_3$ reactor for HCl removal, an HDS reactor for COS/thiophene and two types of H_2S removal (bulk & ZnO) for deep H_2S removal. Because they could all work at temperatures higher than 350°C, it seemed possible to place them in one row before OLGA (Figure 13) or later behind OLGA (Figure 14).

In the final design (Figure 15), this is not the case anymore, because the HDS reactor has to work at a minimum pressure of 3 bar (Martin 't Hoff, Royal Dahlman, personal communication). With the high water content in the gas, increasing the pressure will lead to water condensation in the compressor, which will cause problems in the system. Therefore, we decided that a water removal system is required and thus a water scrubber was included in the final system design. This part of equipment reduces the water content of the gas to about 6% and also dissolves about 98% of HCl plus around 40% of NH_3 (Martin 't Hoff, Royal Dahlman, personal communication). So although the addition of a water scrubber to the system means cooling down the gas even more and results in waste water treatment[10], it is inevitable when the HDS reactor has to be part of the system. As was made clear in 2.3.4 the HDS unit is necessary to convert both thiophene and COS. As an extra advantage, it removes HCN and converts double bonded hydrocarbons, making it a very useful piece of equipment.

Because HCl is now removed by the water scrubber, the HCl reactor was removed from the final design. Because it is somewhat unclear which working temperatures the ZnO reactor can handle (see Table 17), we added a heat exchanger between the HDS and ZnO. The double removal of H_2S was unnecessary at this scale, because the ZnO reactor contains an extra guard bed (Martin 't Hoff, personal communication). This guard bed prevents contamination of the gas with H_2S when the first bed has a breakthrough.

Table 17 Reported reactor temperatures for ZnO reactor

Reactor Temperature	°C	comment
Zuber et al. (2014)	300	<1 ppm H2S
van Paassen et al. (2006)	450	about 0.1 ppm H2S
Zwart et al. (2006)	300	<0.3 ppm
Broust (2008)	400	
van der Meijden (2010)	250	+ extra bed 160
Bocci et al. (2014)	450	max
Nagel (2008)	350-450	
Aravind et al. (2012)	300-550	
Rabou & Almansa Aranda (2015)	200	<0.1ppm (all S)
Average	358.33	

Pre-reformer
The pre-reformer for conversion of higher hydrocarbons is not included in the final system (Figure 15). The earlier designs included a reformer for hydrocarbons higher than methane, such as used in Rabou & Almansa Aranda (2015) for the production of bio-synthetic natural gas (bio-SNG). Eventually, it is considered unnecessary because commercial fuel cells work on natural gas (Ammermann et al., 2015; Mcphail et al., 2013), which also contains small percentages of higher hydrocarbons. Therefore, an SOFC fuel cell system already includes a pre-reformer when working on 100% natural gas to convert 25% of the methane to hydrogen (James et al., 2012). The other 75% of the gas is internally reformed in the fuel cell itself (by nickel catalyst at the anode).

The second reason is that the fuel inlet gas from gasification of biomass methane makes up a minor part of the gas. For example the thesis by Van der Meijden (2010) reports 8.1 vol% of methane in the gas (wet) and 12.9% (dry) after gasification of wood, plus about 3% higher hydrocarbons. This is far less than in natural gas, so an extra pre-reforming unit is not needed. Thus after the ZnO reactor only depressurization to fuel cell pressure (1.4 atm) and heating of the gas to fuel cell temperature. By this depressurization some energy could be earned back, but this is not further considered in this research.

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[10] Though the AEB can probably take care of this, it still contributes to the complexity of the system.

Fuel cell system
The SOFC system includes (James et al., 2012):
- A heat exchanger system for fuel and air preheating
- A post-combustor for unused fuel (plus heat exchangers)
- A pre-reformer for reforming methane and higher hydrocarbons in the gas

The temperatures in the system design are adopted from the gasification-fuel cell system from (Nagel, 2008).
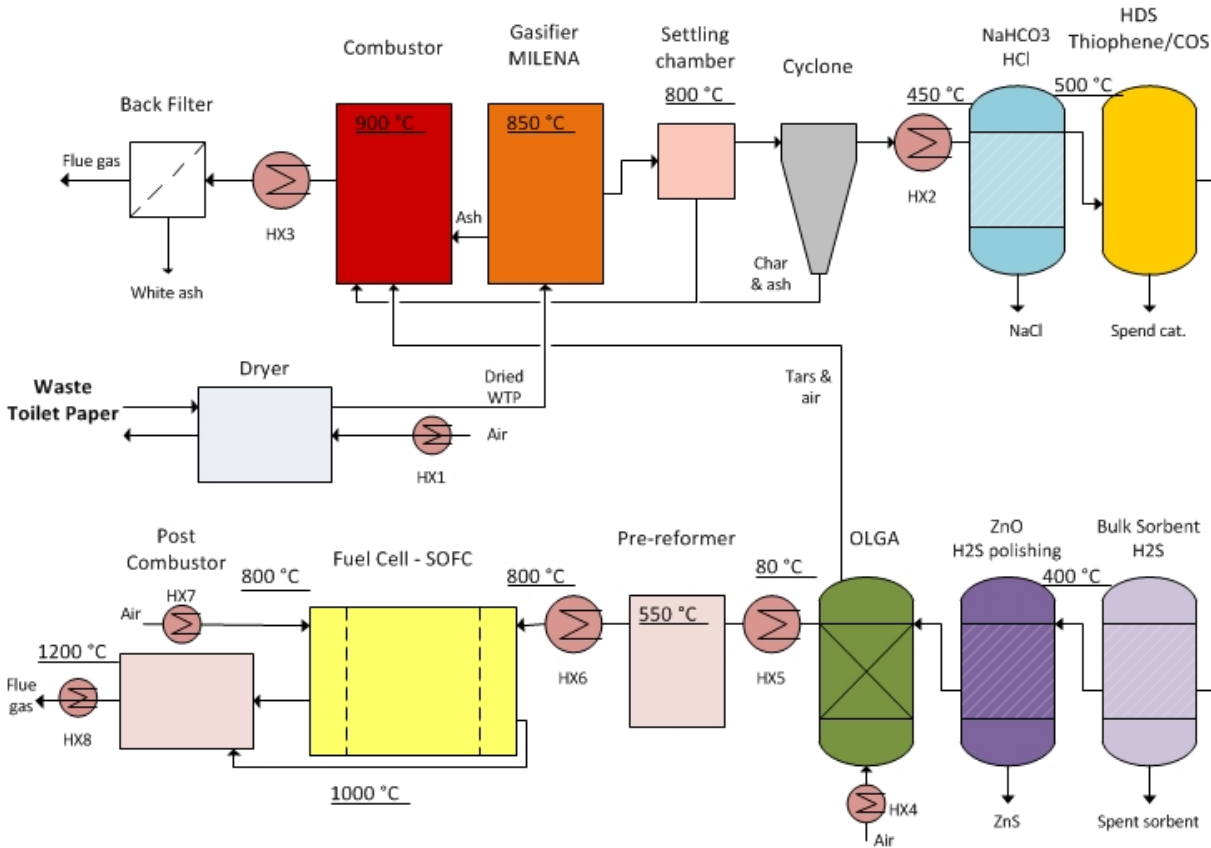
Figure 13 System design, version 1. After meeting with ECN.
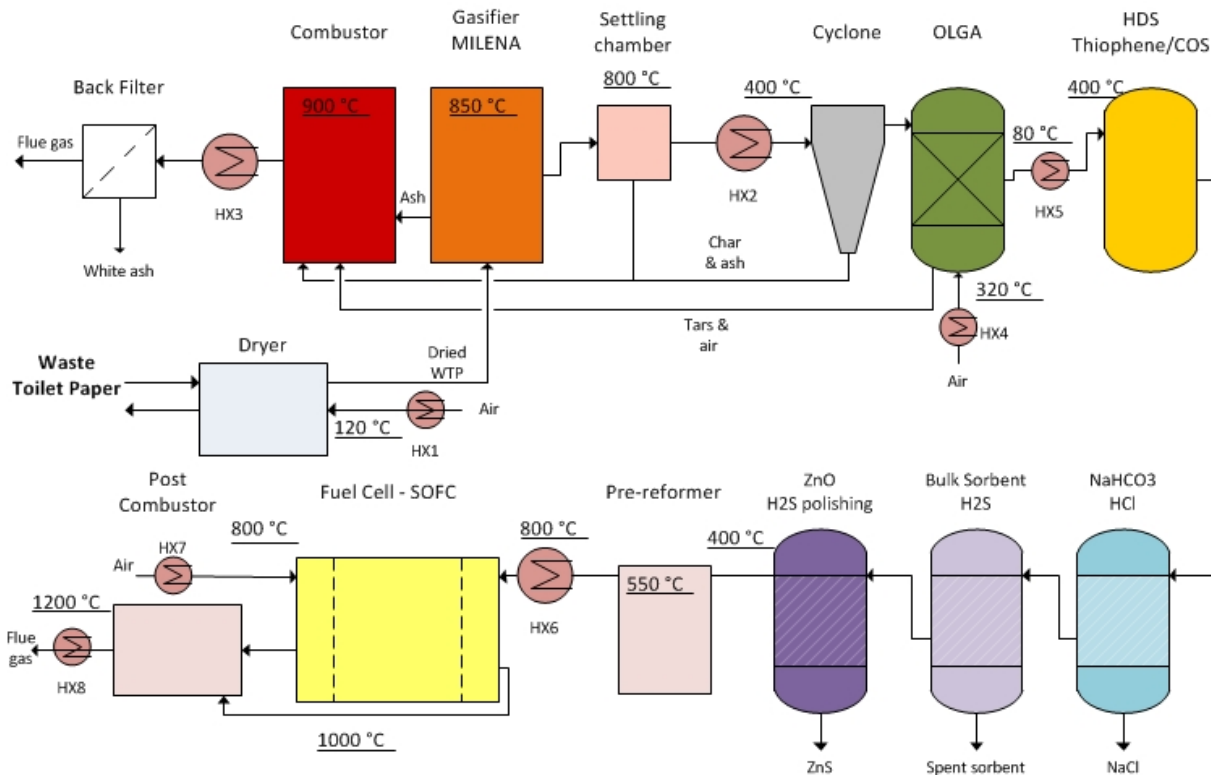
Figure 14 System design, version 2. After email contact with ECN
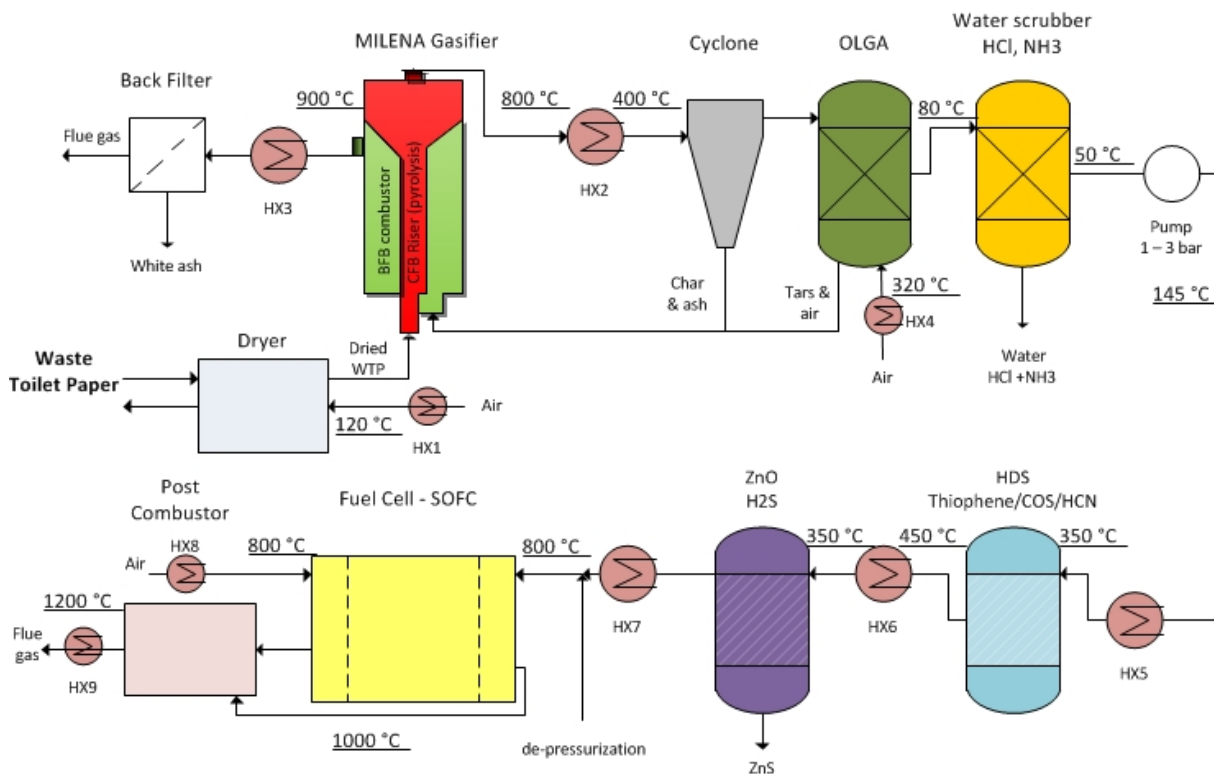
Figure 15 Total system design, final version. After meeting with Royal Dahlman.

4.3. System analysis

4.3.1. Overall energy balance gasifier system

An overall energy balance is made for the system in Table 18, with mass and energy balances of each separate piece of equipment as input, these can be found in Appendix F.

Table 18 Energy balance of the total system, with all energy in and outflows.

Total Energy balance			
In	**dH (kW - LHV)**	**Out**	**dH (kW- LHV)**
WTP	4943	HX2	514
Q sens WTP	28	HX3	1356
Dryer air (HX1)	2410	Flue gas + ash	45
OLGA/gasifier air	18	Combustor heat loss	99
OLGA (HX4) air preheat	480	OLGA heat loss	312
Pump work	94	Water scrubber heat loss	67
HX5	147	Pump heat loss	31
		HX6	170
		ZnO heat loss	14
		Electricity	2890
		Rest heat SOFC	1629
		Heat loss SOFC	970
Sum	8120		8097

4.3.2. Heat integration

To get a first idea of the heat integration possibilities, Table 19 shows the heat sinks and sources in the gasifier and cleaning system. The rejected heat of the fuel cell system is

included. The streams that are matched as heat source and sink have the same colour. So while the air for the OLGA cleaning system is heated up from 15°C to 380°C, the raw product gas from the gasifier can be cooled down from 800°C to 400 °C. The dryer requires such a large amount of heat that two heat sources are needed. It would be most efficient to use the fuel cell flue gas to heat up the dryer air first, and then the combustion flue gas is used to heat up the last part. After heat integration, there is still 573 kW (low temperature heat) left. Of this heat all heat with a temperature of 80°C and higher (555 kW) is assumed to be useful for the district heating system of the AEB.

Table 19 Heat integration options. Streams that have the same colour can be matched.

Heat integration			
Heat sinks	Tin (°C)	Tout (°C)	dH (kW)
Dryer air – HX1	15	120	2410
Olga air – HX3	15	380	480
HDS – HX5	145	350	147
Sum			3036
Heat sources	Tin (°C)	Tout (°C)	dH (kW)
Combustion flue gas – HX3	900	35	1340
Gasification gas – HX2	800	400	514
ZnO cooling – HX6	553	350	170
Fuel cell flue gas – HX9	200		1625
Sum			3650
Rest heat			573
Usable rest heat (> 80°C)			555

4.3.3. Electricity production and total system efficiency

The cleaned product gas from the gasifier is fed to a fuel cell with 55% electrical efficiency (LHV) and 31% efficiency to heat (Ammermann et al., 2015). This means that from the starting material of 25,000 tonne wet WTP/year or 3567 kg/hr with 80% availability, 2.883 MW of electricity can be produced. On a yearly basis this is 20.2 GWh or about 6400 Amsterdam households of 3150 kWh/year (CBS, 2015).

The total system efficiency for producing electricity is given in Table 20. Here the leftover heat (after heat integration) is included as product and the electricity need for the gasifier and cleaning system is deducted from the electricity production. By doing this, an electrical efficiency from WTP (wet) to electricity of 57.2% is obtained. The total efficiency including heat is *ca.* 70%.

Table 20 Total system efficiency

Total system efficiency			
In	kW	**Out**	kW
WTP[1]	4943	Electricity	2825
		Heat	622
Total	*4943*		*3446*
Electrical efficiency		57.2%	
Total efficiency (incl. heat)		69.7%	

[1]This WTP has 60% moisture. At the inlet of the gasifier the WTP is dried to 25% moisture and the energy content (LHV) = 6071 kW – see appendix F. So the MILENA gasfier should have a scale of 6MW, and not 4.9 MW.

4.4. Economic analysis

With input data from 3.6, economic indicators were calculated for the system with and without subsidy. The results of these calculations are shown in Table 21 and Graph 5. Table 21 shows that in both cases the investment cannot reach break-even, which is the reason that no (discounted) payback period is given. The NPV is negative in both cases, which means that the business case is not going to be profitable over the lifetime of the project. The IRR is negative and this indicates a negative rate of return on the investment. The LCOE shows that without subsidy the electricity price is 23.05 eurocent/kWh, which is comparable to the consumer price for electricity. With subsidy the price is 12.67 eurocent/kWh but this is still 7.77 eurocent/kWh higher than the assumed price for electricity production of 4.9 eurocent/kWh. To reach break-even (an NPV of zero), 15.98 eurocent/kWh of subsidy is required.

Table 21 Results of the economic analysis

Results	Original system	With subsidy
Discounted PBP (years)	-	-
NPV (M€)	-38.20	-22.03
IRR (%)	-24.2%	-9.7%
LCOE (eurocent/kWh)	23.05	12.67
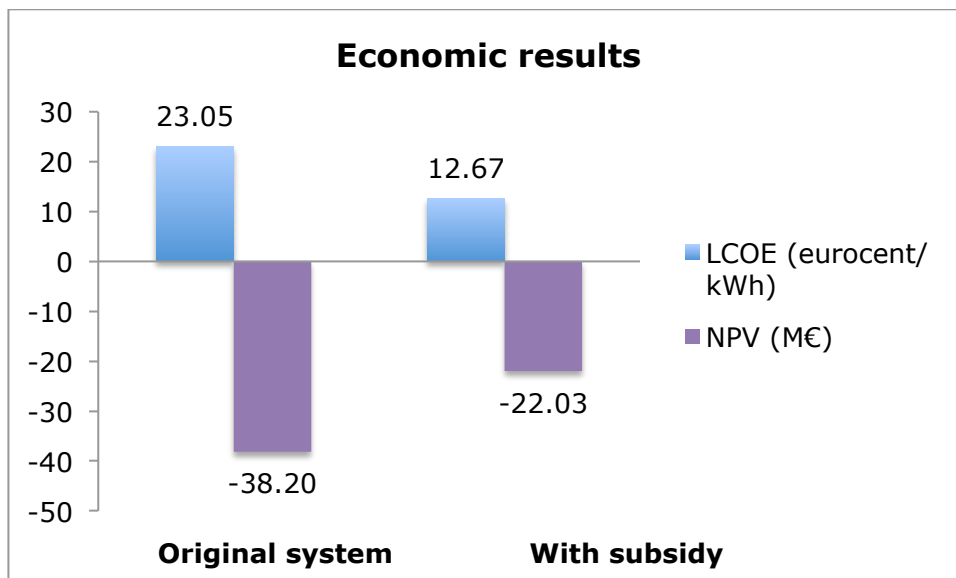
Graph 5 NPV and LCOE of a WTP-to-electricity plant with and without subsidy

4.5. Learning

The effects of learning are limited by the effect that a piece of equipment can have on the LCOE. Therefore, the build-up of the LCOE was determined and shown in Graph 6. The SOFC has a large share of 42% in the LCOE (initial investment plus later stack replacements), while the gasifier makes up 17%. The CAPEX including later SOFC investment make up 80% of the LCOE. When stack replacements for the SOFC are considered as OPEX, the CAPEX still makes up 65% of the LCOE. Thus the LCOE is mostly driven by the CAPEX/initial investments, and to a lesser extent by the OPEX/OM cost.
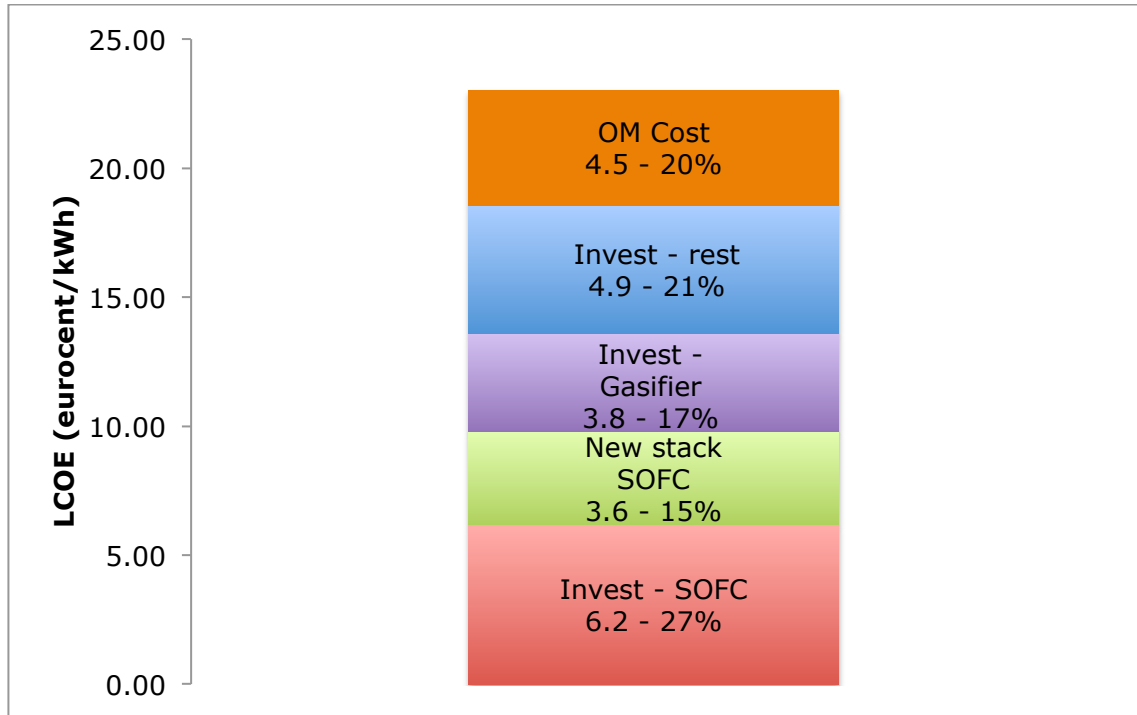
Graph 6 Build-up of LCOE. Invest = investment costs., New stack SOFC contains the SOFC investments that have to be done during the project lifetime.

With Eq 11 & 12 the effect of learning of the gasifier system and SOFC on LCOE was investigated. The global cumulative shipments[11] of SOFC systems is still limited with 194 MW, but when the industry develops and more SOFC systems are produced, the costs for SOFC systems will decrease. In Graph 7 the effects of learning are shown when the cumulative global shipments for SOFC will increase from the current 0.194 GW to 150 GW. The learning effect is stabilizing around 15 cents/kWh after 100GW cumulative shipments. At this price, the NPV of the system is -18.5 Meuro, so the plant would reach a break-even. When subsidy is added (here assumed to be constant) the LCOE comes close to the average market price for electricity production at 5.14 eurocent/kWh, but the NPV is still negative at -2.3 Meuro.

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[11] This parameter is used as a proxy for the total installed capacity, see for more information appendix H.
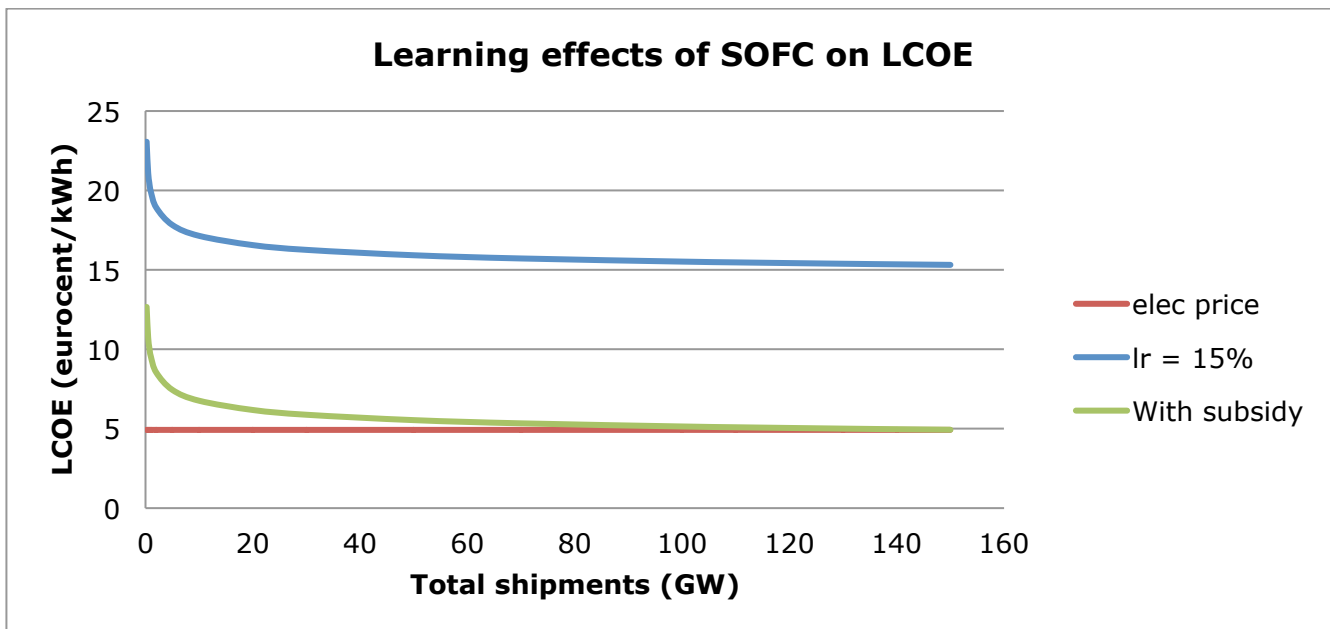
Graph 7 The effect of learning in the SOFC market on the LCOE, both initial investments plus later stack replacements are included. The red line shows the current average price for electricity production in the Netherlands over the past 10 years (APX, 2016). The blue line shows the LCOE development with a SOFC learning rate of 15%. The green line shows the combined effects of learning (15% learning rate) and subsidy (10.38 eurocent/kWh).

The gasifier system cost has less impact on the LCOE than the SOFC, which is also clear from the result in Graph 8. With only learning effects from the gasifier, the cost of electricity production will only decrease just below the 20 eurocent/kWh, also when the total installed capacity of biomass gasifiers will increase vastly. After the first 150GW, the learning effect reaches a plateau.
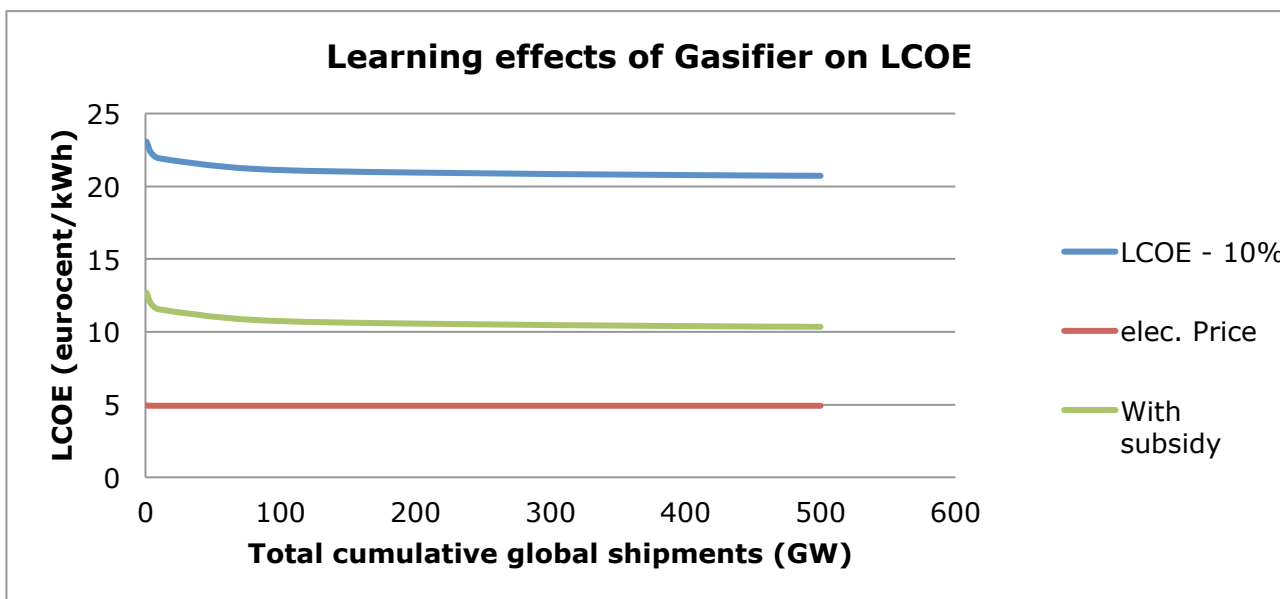
Graph 8 The effect of learning in gasifier system on LCOE. Blue: learning rate of 10% on the investment costs of the gasifier system. Green: with 10.38 eurocent/kWh subsidy. Red: the current average price for electricity production in the Netherlands over the past 10 years (APX, 2016).

4.6. Uncertainty

4.6.1. Results sensitivity analysis

For the parameters that could be combined in one graph, the sensitivity results are shown in Graph 8. The x-axis is given as percentage of deviation from the original value (see Table 8) in order to combine the results for different parameters in one graph. For other parameters the results are shown separately. The complete results including all data tables can be found in appendix I.
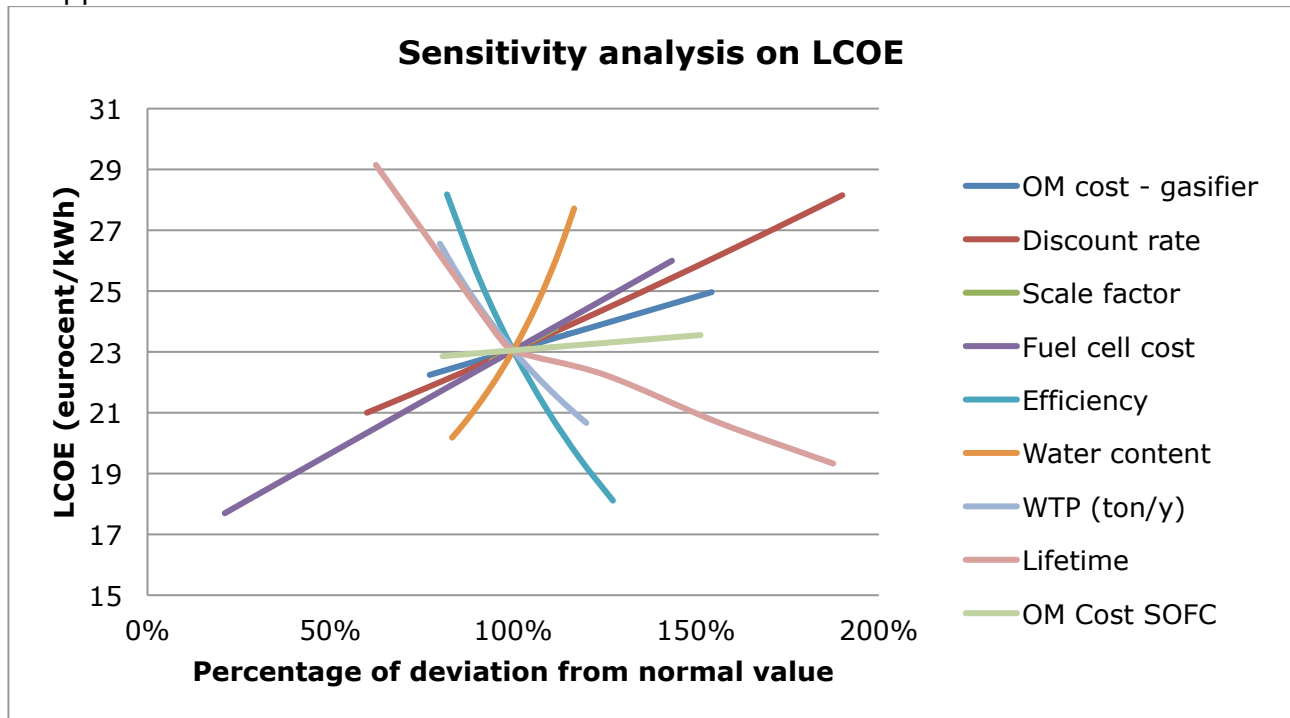
Graph 9 Results of sensitivity analysis for OM cost, discount rate, scale factor, fuel cell cost, fuel cell efficiency, water content in WTP and amount of WTP.

WTP price/Fuel price

Because the initial WTP price is set at 0€/tonne, we could not give the percentage of deviation of this parameter in such a way that the results could be included in Graph 9. Therefore, the results are shown individually in Graph 10. The graph shows a straight line, indicating that when the costs of generating waste increase, the economic feasibility of the system improves.
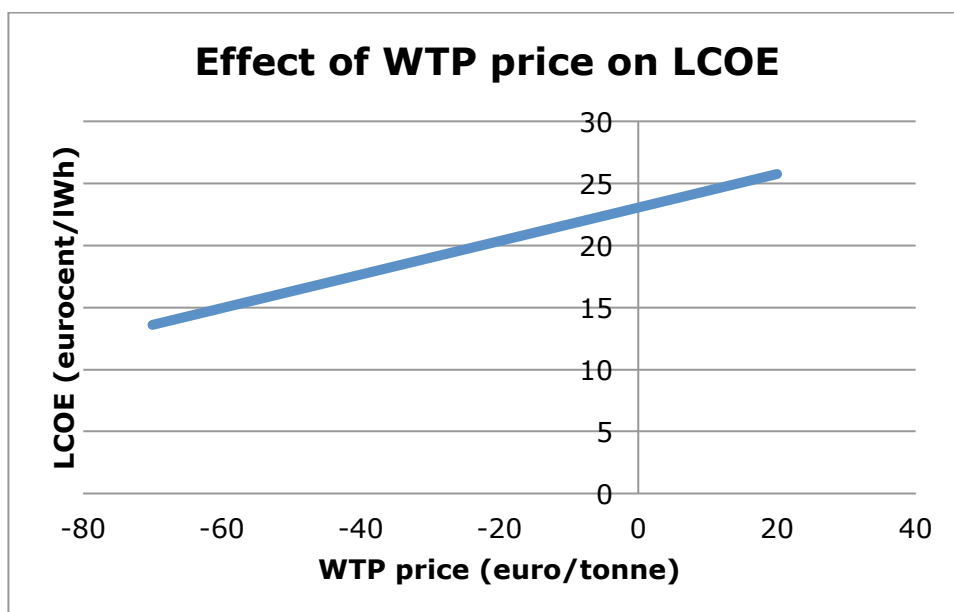
Graph 10 Sensitivity of WTP price on LCOE

Learning rate SOFC

A graph with different learning rates was created to look at the effect on the LCOE (Graph 11). As different learning rates for the gasifier system have almost no effect they are not shown here. When the learning rate increases, the LCOE becomes lower, though this effect is limited. The difference between 10% and 15% learning curves is larger than the difference between the 20% an 25% learning curve.
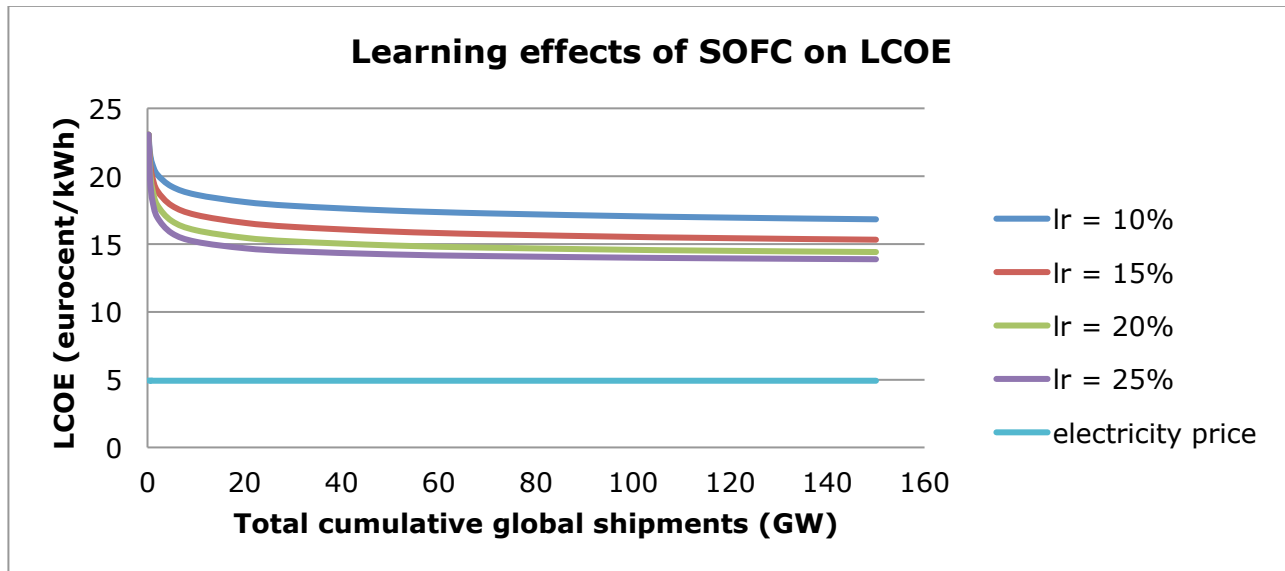
Graph 11 Sensitivity of SOFC learning rate on LCOE, both initial investments plus later stack replacements are included. The upper lines show the LCOE development under different SOFC learning rates. The lowest (light blue) line shows the current average price for electricity production in the Netherlands over the past 10 years (APX, 2016).

WTP composition

Two different WTP compositions were used as input for the gasification model (sample 2 & sample 3 from Table 10) and the results on the LCOE were 22.69 eurocent/kWh for the first variation and 23.22 €cents/kWh for the second variation.

Summary of the results

In order to compare all results from the sensitivity analysis, Table 22 was created. This table shows the average deviation of a parameter from the standard value (23.05 eurocent/kWh). In the final column the values are normalized to the highest value, subsidy value. The table shows that other sensitive parameters are price of WTP, efficiency of the SOFC and project lifetime.

Table 22 Sensitivity values, the maximum decrease and increase from the centre value (23.05 cents/kWh) is calculated and divided by two in the second column. In the third column the results are normalized to the largest value (7.50 – subsidy value).

	Sensitivity (average deviation from LCOE in eurocent/kWh)	Normalized
Amount of WTP	2.95	0.39
Water content WTP	3.77	0.50
WTP composition	0.26	0.04
Price of WTP	7.43	0.99
SOFC investment costs	4.15	0.55
Efficiency SOFC	5.03	0.67
Scale factor SOFC	0.55	0.07
Learning rate SOFC	1.53	0.20
Project Lifetime	4.91	0.66
OM costs- Gasifier	1.36	0.18
OM cost - SOFC	0.35	0.05
Subsidy value	7.50	1.00
Discount rate	3.88	0.52

4.6.2. Results pedigree analysis

The results from the pedigree-analysis from five experts plus the author are summarized in Table 23. All medians have an input of a minimum of four scores, because not all experts assigned scores to each parameter. The separate pedigree-matrices of each expert can be found in Appendix J. Table 23 shows a clear division the scores for different pedigree criteria. Overall the parameters have good (>3) scores on proxy, but low scores on validation process. Reliability of source is clearly low for the SOFC parameters and higher for the general project parameters (project lifetime, OM cost gasifier, subsidy and discount rate). The standard deviations give some insight in how the score varied among the different experts.

Table 23 Summary of pedigree-analysis

Pedigree Criteria / Input parameters	Proxy Median	Stdev	Reliability of source Median	Stdev	Completeness Median	Stdev	Validation process Median	Stdev
Amount of toilet paper	3.5	0.82	1.5	1.51	0.0	0.84	1.0	0.71
Water content toilet paper	4.0	0.89	1.0	0.84	1.0	1.14	1.0	0.45
WTP composition	3.0	0.84	4.0	1.30	3.0	1.14	1.0	1.00
Price of WTP	4.0	0.89	3.0	1.64	2.0	1.10	0.0	1.41
SOFC investment costs	2.5	0.58	1.5	0.58	2.0	1.00	2.0	1.50
Efficiency SOFC	3.0	0.50	1.5	0.58	2.5	0.58	2.0	0.00
Scale factor SOFC	3.0	0.82	1.0	0.82	0.5	0.96	0.0	0.00
Learning rate SOFC	3.0	0.50	2.0	0.00	2.0	1.15	2.5	0.96
Project Lifetime	4.0	0.89	2.0	1.00	2.0	1.22	2.0	0.84
OM costs- Gasifier	3.0	0.55	4.0	0.45	3.0	0.45	2.0	0.71
OM cost - SOFC	3.0	0.82	1.0	0.50	1.5	0.58	0.5	0.58
Subsidy value	2.0	0.84	4.0	1.64	2.0	0.89	2.0	0.89
Discount rate	4.0	0.55	3.5	1.41	2.5	1.29	3.0	0.00

4.6.3. Results diagnostic diagram

The input data from Table 22 and Table 23 were combined in a diagnostic diagram. The diagram is divided into four quadrants; low pedigree/high sensitivity (red), low pedigree/low sensitivity (yellow), high pedigree, high sensitivity (yellow), high pedigree, low sensitivity (green). In Graph 12 the 13 input parameters are shown as points with as x-axes value the pedigree score and as y-axes value the normalized sensitivity.

The SOFC Investment cost and the water content of WTP are both in the red quadrant, which means they have the highest priority in further research as they do have a relatively high impact on the LCOE, but a low pedigree. The next priority could be on the amount of WTP, as its sensitivity is near the red quadrant while its strength is also low. The other parameters in the lower left quadrant have a minimal impact on the LCOE, also when their values would change when more knowledge is gained. From the other parameters with a pedigree higher than two, the strength is high enough. Though, the parameters in the upper right quadrant do have a higher sensitivity than the ones in the lower right quadrant, which is the reason their colour is yellow instead of green.
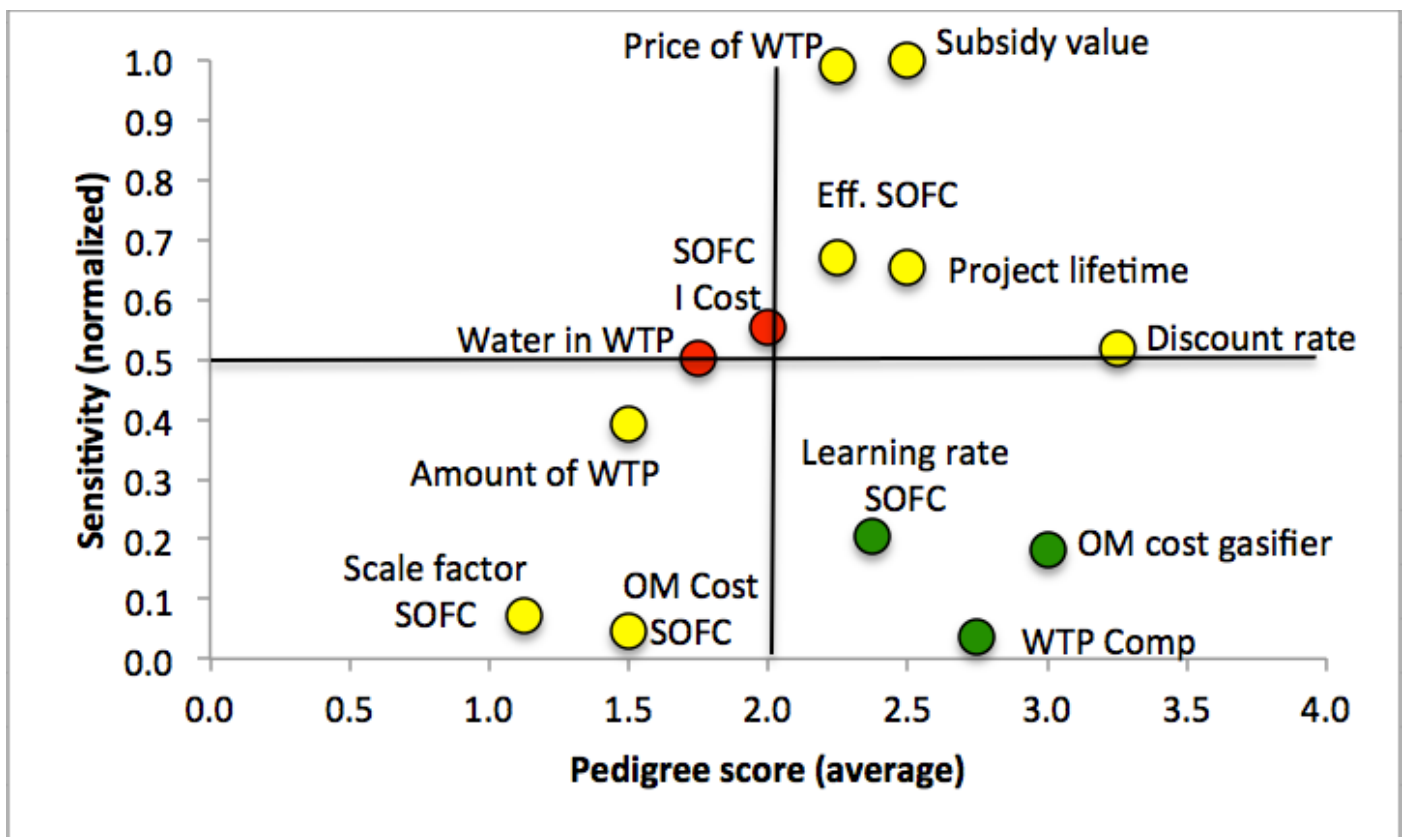
Graph 12 Diagnostic diagram. The x-axes shows the strength of the parameters as average of the pedigree score. The y-axes shows the normalized sensitivity of the parameters on the LCOE.

5. Discussion

Now the results are given, it is time to reflect on and analyse their meaning, discuss discrepancies and compare the results with other meaningful data. Reflection and analysis will be done pointwise in the first section. The discussion of the results is followed by the discussion of the method, to point out the limitations of the method used.

5.1.   Discussion of the results

- *Energy balance:* The gasifier heat balance (Table 35) is not solved, about 1200 kW of heat is missing. Though no extra heat was added to the gasifier system because the heat balance of the total system is solved, see Table 18. Still it is interesting to think of reasons that explain why with the same assumptions as in the wood gasification model, heat is missing. The first possible reason is that WTP has a lower energy content than wood. In van der Meijdens' model (2010) wood has an energy content of 18.5 MJ/kg LHV and 19.9 MJ/kg HHV while WTP has an energy content of 17.49 (HHV) and 16.13 (LHV), both dried. So it could be that extra heat is just necessary in this system.

    Another reason could be that during the modelling we allocated too much latent heat to the flue gas, as more heat in the flue gas means less heat available to heat up the gasifier. The inlet of WTP in our model is less than wood in the model of van der Meijden (about 1902 kg/hr WTP vs 2717 kg/hr wet wood), but more air is needed in the combustor with a lambda of 1.3 (4444kg/hr for WTP instead of 3623 for wood), which also means that a larger volume of flue gas is created (4578 kg/hr for WTP gasification vs 3734 kg/hr for wood gasification).

Energy efficiency: The total electric efficiency is 57.2%, which is quite high when compared to other electricity generation technologies as is done in Graph 13. Natural gas combined cycle systems have a comparable electrical efficiency, but are a really different category. The most likely alternative for WTP conversion would be waste incineration, with a twice as low electrical efficiency. Compared to similar systems that work with biomass gasifiers with an SOFC – B-IGSOFC - this study shows higher electrical efficiency. This could be due to the more recent data that were used in this study, but in any case it shows that gasification – fuel cell systems are promising in terms of electrical efficiency.
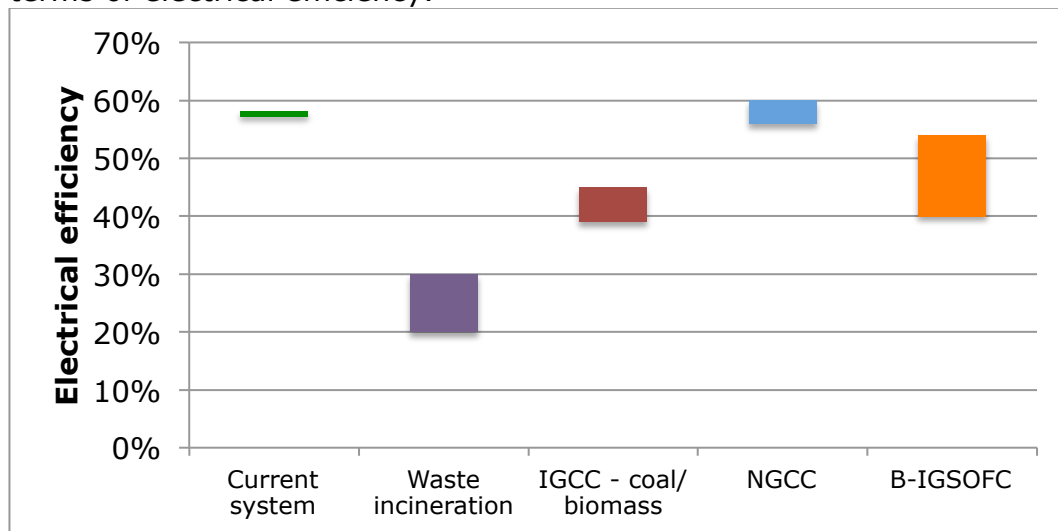
Graph 13 Comparison of WTP to electricity with other electricity production technologies. Green is the current system as discussed in this report. Purple is waste incineration, with an electrical efficiency varying between 20-30% (European Commission, 2006; Gemeente Amsterdam & AEB, 2012). Red is Integrated Gasification combined cycle for a combination of coal and biomass (van den Broek et al., 2009), Blue is Natural Gas Combined Cycle (van den Broek et al., 2009; Graus et al., 2008; Thattai et al., 2014) and orange shows the efficiency range of different scales of Integrated biomass gasification SOFC – Gas turbine (B-IGSOFC) – some wit gas turbine- systems (Aravind et al., 2009; Jin et al., 2009; Nagel et al., 2009).

- *Economic analysis – CAPEX vs OPEX:* At the moment the OM costs for the gasifier system are based on a stand-alone system according to the information from Royal Dahlman. When the system would be built at the AEB site, the labour costs could decrease significantly as employees could be shared between installations. Though, the LCOE build up in Graph 6 shows that the influence of lower labour costs will be minor, as the fuel cell investment costs take up the largest part of the LCOE. Fuel cell price is therefore an important factor in learning effects, and we showed that when the SOFC market grows from the current 194 MW to 50 GW, the LCOE will decrease from 23 to 15 cents/kWh. Yet, this is still not enough to make the system break-even.

  Therefore an additional scenario is considered in which the initial investments for the system are excluded from the analysis, which means that only the OPEX cost influence the feasibility of the system. The capital investment is then considered as something that can be paid off as a loss when the investment is of benefit to society. Two scenarios are analysed. The first scenario includes both OPEX and the stack replacements for the fuel cell that have to be done during the project lifetime. The second scenario includes only OPEX costs. Results are shown in Graph 14, and make clear that the system becomes feasible when only the OPEX cost are taken into account.
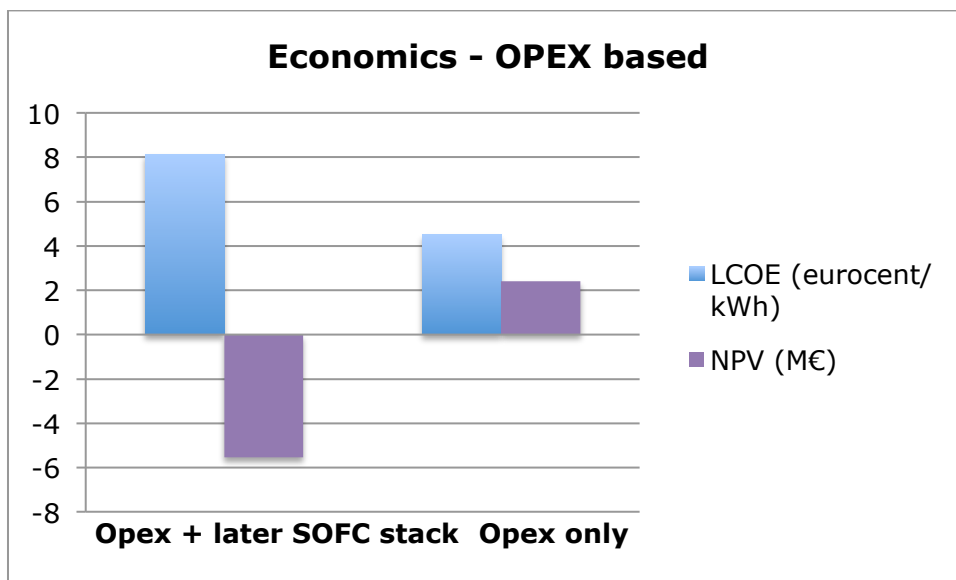
Graph 14 The NPV and LCOE of two OPEX-only scenarios. The left part of the graph shows the results when both OPEX and stack replacement costs for the SOFC are considered. The right side shows the results with only OPEX cost.

- *Economic analysis – Incineration scenarios:* Because the system is not economically feasible, an alternative idea is to build the whole system except for the fuel cell and use the output gas as feed for the waste incinerator of the AEB. The AEB has a waste incineration system with an electrical efficiency of around 30% (Gemeente Amsterdam & AEB, 2012). This is an option until the fuel cell becomes cheaper, as the sensitivity analysis and LCOE build up showed that fuel cell factors have a large effect on the LCOE. Another option could be to only dry the WTP and use it as input for the waste incinerator, though then the flue gas has to be thoroughly cleaned afterwards.

  For both cases, an economic analysis was done. An electric efficiency of 30% for the input was assumed; extra heat needed in the system lowers the overall electric efficiency. For the economic analysis the investment and OM costs of the removed parts of the system were excluded from the calculations. Costs for heat were set at 5 euro/GJ (the same price as for heat production).

  The electricity produced with drying and incineration is 1.8 MW, with the whole system set up except the fuel cell and including incineration the electricity production is 1.6 MW, versus 2.8MW with the original system. Graph 15 shows the economic results for both scenarios. The total electrical efficiency of the systems are lower then for

original system (24 and 25% vs 57%), but the economics are more promising. When the WTP is only dried and then incinerated, the project even becomes profitable with an NPV of 6.35 million euro. For a system that is almost complete except for the fuel cell, the NPV is still negative and the LCOE 20 cents/kWh instead of 23.05 cents/kWh for the full system. The small difference with the original sytem is due to the much lower electricity production by incineration.

Economics - Incineration scenarios

Legend:
- ■ LCOE (eurocent/kWh)
- ■ NPV (M€)

Values shown: Drying + Incineration: 0.80, 6.35; Gasify, clean + incin.: 19.98, -20.11
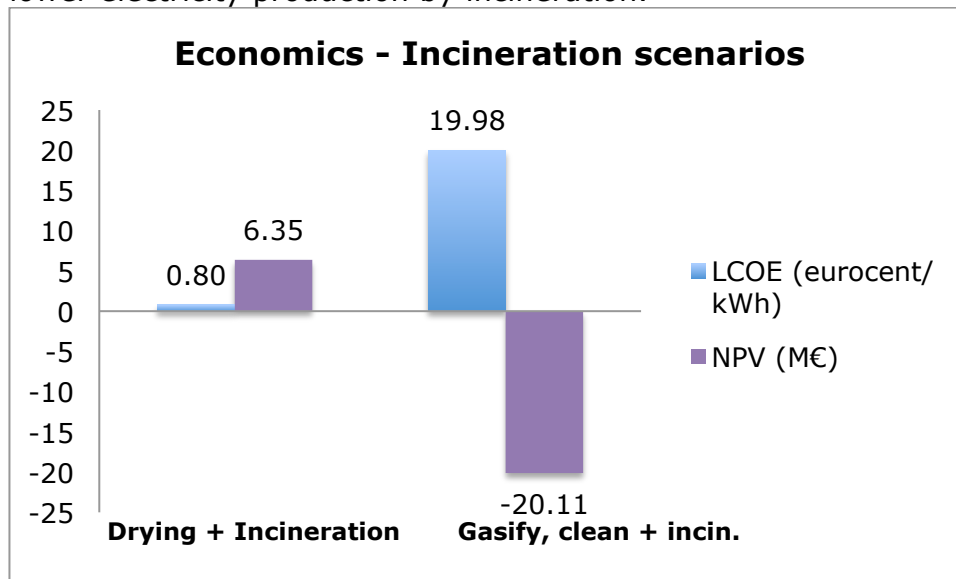
Graph 15 The NPV and LCOE for a system with only drying of WTP followed by incineration (left), and a whole system set up with gasifier and cleaning system followed by incineration, but without SOFC.

- *Economic analysis – comparison with renewable technologies:* Producing electricity from waste is not the same as from renewables. Though, since toilet paper is made from trees, which are in the short carbon cycle, WTP could also be considered as biomass. Therefore, it might be interesting to compare the LCOE derived in this study to the LCOE of renewable energy technologies. In Graph 16 the LCOE of the current system (23.05 eurocent/kWh) is compared to electricity from PV, wind and biogas. The WTP to electricity system is most similar to electricity production from biogas, and also shows the smallest cost difference with this technology. Though, it is clear that the WTP to electricity system is not able to compete with electricity from these renewable technologies, as the LCOE is significantly higher.
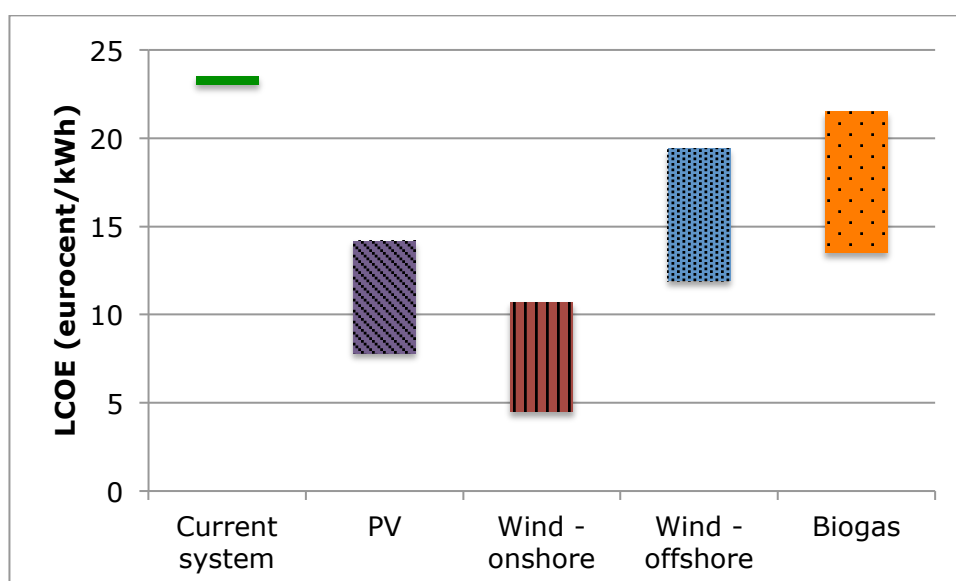
Graph 16 Comparison of LCOE to renewable energy technologies from (Kost et al., 2013).

- *Pedigree analysis:* The pedigree workshop results showed overall low (<2) scores for validation process (see Table 23). This is not too surprising as we did a first stage feasibility study. The concept of converting WTP into electricity is new, which makes validation with existing plants impossible. In addition, it is also clear that the experts do not agree with each other on the scores of some parameters as can be concluded from the high standard deviations for these parameters. The variation in parameter scores may be due to the low number of experts and thus scores (max 6 scores per parameter/indicator combination) yet the perspectives of the different experts vary as well.

    Experts from Waternet and Royal Dalhman have a more in-depth view on some parameters than the information that was provided in the information document. For example someone from Waternet might know that the starting material for the ultimate analysis (WTP composition) is not 100% representative, while other people will just read that material is obtained from Waternet (and will think this is a reliable source) and only assess the method of the ultimate analysis itself. This makes the results less consistent, but also more interesting.

- *Pedigree analysis:* Discussion during the pedigree analysis revealed that Royal Dahlman considers systems under 20MW unfeasible (without subsidy). This would mean that the scale of the installation has to be increased fourfold. This could be done by combining WTP with other types of cellulose waste from the AEB, but it is questionable if the gasifier would be able to handle this. Another option would be to collect WTP from different WWTP in the province of Noord-Holland and parts of Utrecht and/or Zuid- Holland. The drawback here would be the transport costs.

5.2. Discussion of the methodology

- *Gasification model:* The main assumption in the gasification model is that WTP will behave similar to wood during the gasification process. This main assumption induced other simplifications and assumptions such as:
    o The multiplication factors for biomass/WTP to mass product gas, char and tar.
    o The composition of tar and char

Because these assumptions were not checked experimentally, it is hard to judge how reliable the outcomes of the product gas composition are.

    The main difference between WTP and wood is the cellulose content. WTP contains about 70-80% cellulose, while wood has 40-80% cellulose, and also hemicellulose (15-30%) and lignin (25%-35%) (Nagel, 2008). The structure of hemicellulose and lignin are given in Figure 16. Cellulose contains relatively more oxygen than carbon, which also explains that no extra oxygen was needed as gasifier input in the analysed system.
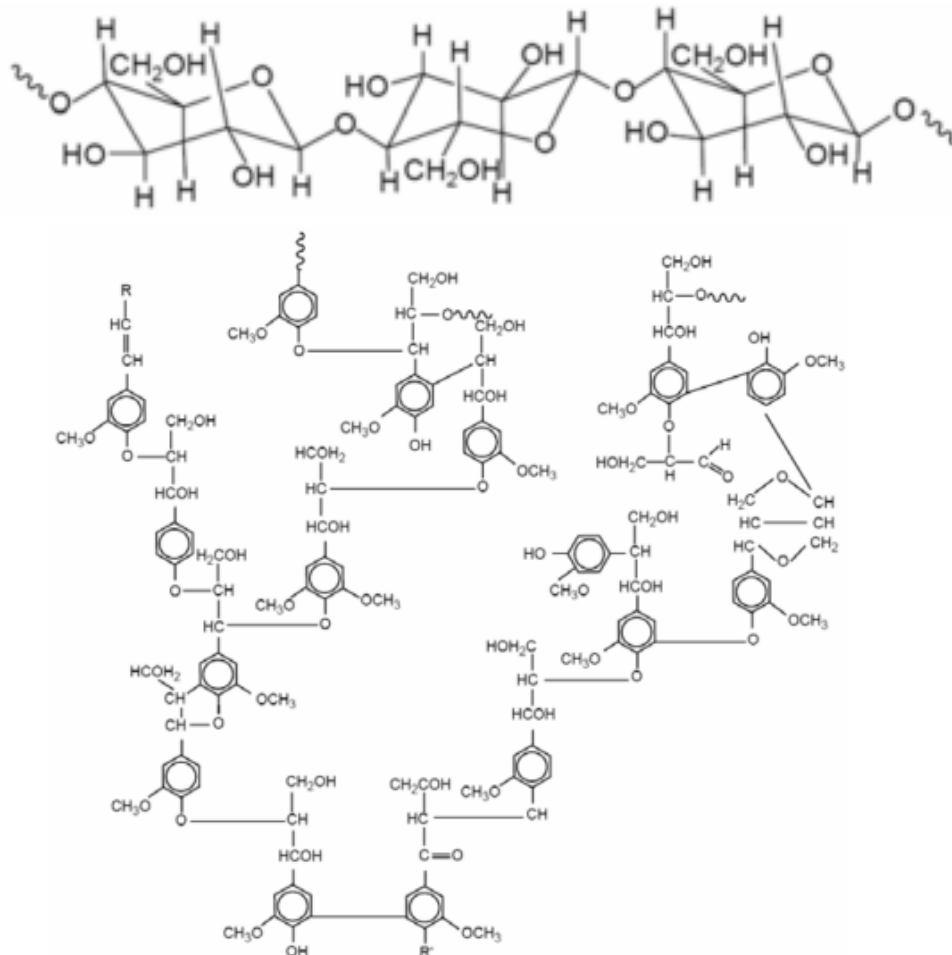
Figure 16 The structure of cellulose and lignin (Nagel, 2008)

Other effects on the gasification process are a decrease in char and tar yield with increased cellulose, but an higher biomass to gas conversion rate (Lv et al., 2010). The high biomass to gas conversion rate was also confirmed by Wu et al. (2013) who showed that cellulose gasification results in the highest biomass conversion rate to gas (55.3% wt.) and more CO and H_2 in the product gas with cellulose than hemicellulose. Cellulose will still form some higher hydrocarbons, but significantly less than lignin (Wu et al, 2013). Based on these papers the hypothesis is that gasifcation of WTP would give more product gas with relatively more CO, but less char and tar. Because char and tar are fuels, these lower quantities could result in required extra heat input for the gasifier.

- *System analysis:* Heat integration was done in a simplified way, for a more accurate idea of the heat integration possibilities; a proper pinch analysis should be done.

- *Economic analysis:* SOFC costs were derived from general reports and scaled with an unvalidated scale factor. Therefore it would be more accurate to obtain quotations from SOFC suppliers.

- *Learning:* In the learning curve analysis, both the selling price for electricity and the subsidy value were held constant. In reality, it is expected that both electricity prices and subsidies will change over time as the SOFC market develops. It is thus recommended to include a prediction of future electricity (and subsidy) in a more detailed analysis.

- *Pedigree analysis:* The pedigree analysis was done with 6 people, and only four of them assigned scores to all parameters. Furthermore, the experts from the companies did not have experience with pedigree analysis, while some of the people with experience on pedigree analysis might not be considered as experts in the field and

therefore had to base their scores solely on the information given. This may have lead to less reliable scoring results and could explain the sometimes large difference in scores from the participants. When more and better-informed experts fill in the pedigree matrix, the outcomes are likely to be more reliable.

- *Pedigree analysis:* Later evaluation of the pedigree workshop with a supervisor pointed out that the information document that was used during the pedigree workshop sometimes contained some sentences that expressed an opinion. This should not happen as it may point the expert in a certain direction.

6. Conclusions and recommendations

The goal of this study was to assess the techno-economic feasibility of a waste toilet paper to electricity plant. In summary, we state that the goal of the research is reached. This conclusion is supported by the answers on the research questions formulated at the start of the research, which are given below. Aftewards, the final conclusions and recommendations are given, based on a company-, government- and research perspective.

6.1. Answers to research questions

6.1.1. What could be a possible process design for a WTP to electricity plant?

The final process design is given in Figure 15 and includes a dryer, MILENA gasifier, cyclone, OLGA tar removal system, flue gas filter, water scrubber, HDS reactor, ZnO reactor and SOFC fuel cell, plus a network of heat exchangers and a compressor and depressurization unit. With this design, the WTP with 60% moisture can be gasified, cleaned to SOFC inlet requirements and be converted to electricity.

6.1.2. How would such a process perform in terms of energy efficiency and electricity yield?

The electricity production with 5MWth input of WTP will be 2.8MW, or 20.2 GWh, which is enough for 6400 Amsterdam households. The energy efficiency of the total process is 69.7% and the electricity efficiency is 57.2%. This is a high efficiency when compared waste incineration (max 30% electrical efficiency) or other biomass gasification – SOFC systems, as can be seen in Graph 13.

6.1.3. What is the economic feasibility of the process?

The NPV for the system is -38.20 million euro, with an IRR of -24.2% and an LCOE of 23.05 eurocent/kWh, the payback period could not be calculated. These results make the system economically unfeasible. With subsidy of 10.38 eurocent/kWh the system can still not reach break-even with an NPV of -22.03 Meuro, an IRR of -9.7%, and an LCOE of 12.67 eurocents/kWh.

The investment costs make up the largest part of the LCOE as can be seen in Graph 6: 80% including later SOFC stack replacements and 65% without later stack replacements. The largest investment is the SOFC system, which makes up 27% of the LCOE only by initial investment, and even 42% with both initial investment and later stack replacements.

To investigate the economics with less influence of investments in SOFC and other components, the NPV and LCOE were calculated for three pairs of scenarios. In one pair the SOFC costs have decreased because of learning, in another pair all initial investments have been removed from the analysis and in a final pair only part of the system is built followed by incineration. All results are summarized in Graph 17. The main conclusions are that SOFC learning alone will not make the system feasible, even when subsidy is included. A system with only OPEX costs will be profitable, but not when later stack replacements for SOFC are included. If only part of the installation would be build and combined with the incinerator at the AEB site, the most profitable option would be to only dry the WTP to 25% moisture. Thus, the original system can only become feasible when all initial investments including stack replacements are considered as sunk costs. Though, the most profitable way to convert the waste toilet paper is to just dry and incinerate it.
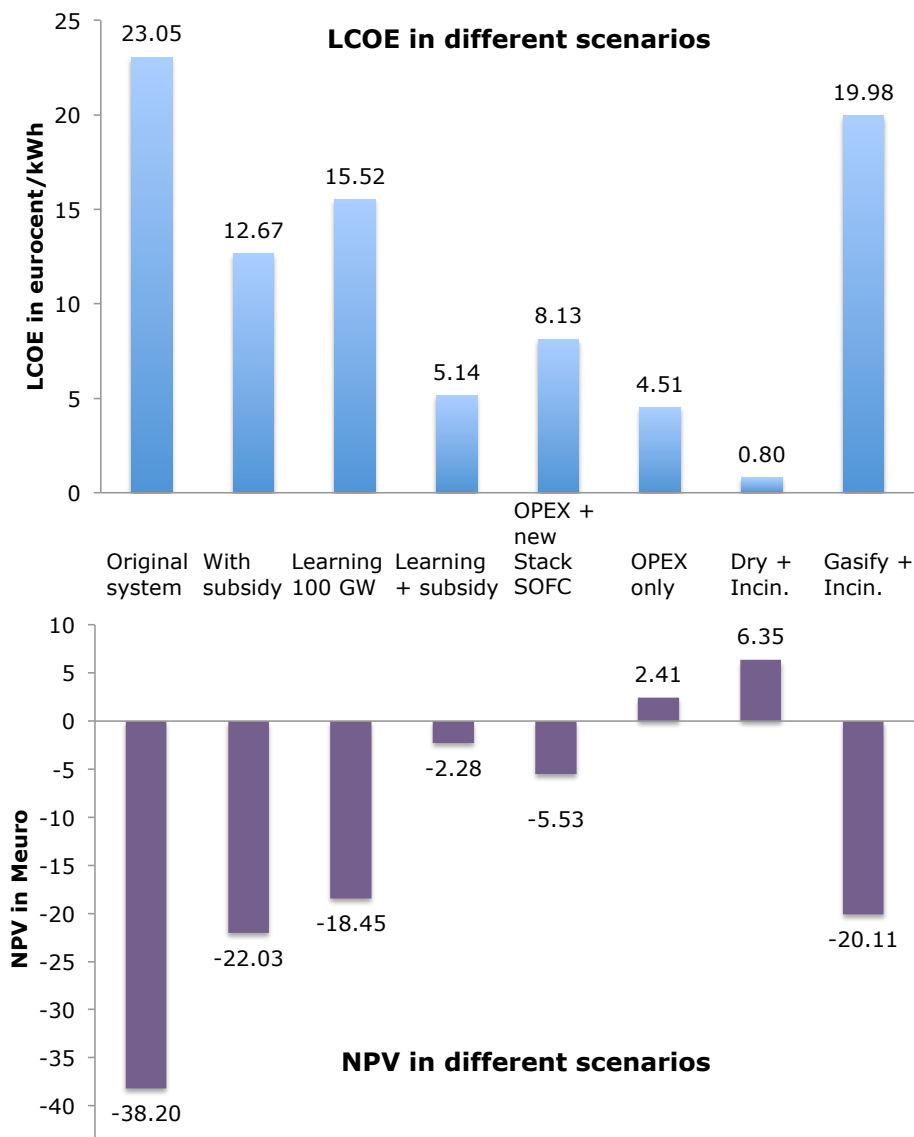
Graph 17 LCOE (above) and NPV (below) in different scenarios. The original system and system with subsidy. Then the system when the SOFC market has grown to 100GW, with and without subsidy. Next two systems with only OPEX and new SOFC stacks or OPEX only. Then finally two incineration scenarios, one with only WTP drying followed by incineration, the other with drying, gasification and cleaning followed by incineration.

6.1.4. What are the largest sources of uncertainty in the results?

The combination of a sensitivity and pedigree analysis resulted in a diagnostic diagram that can be found in Graph 12. This diagram helps to identify the largest sources of uncertainty. These are mainly the SOFC investment cost, water content in WTP and amount of WTP because their impact on the LCOE is high while their strength is low.

Another large source of uncertainty that was not dealt with during the sensitivity analysis is the product gas composition. The excel model that was developed to calculate the gas composition with WTP as input was based on wood. It is expected that gasification of WTP will give different results compared to wood due to the higher cellulose content, but this was not verified and therefore causes an unknown amount of uncertainty in the results.

6.2.   Overall conclusion and recommendations

This study was based on a specific case for the water company Waternet and the waste incineration company AEB and therefore an overall conclusion and advice is given based on their interests. However, the relevance of this study can also be seen in a broader social perspective or from the viewpoint of research. Therefore, the final conclusions and recommendations are divided in three parts, each with a slightly different emphasis.

6.2.1. AEB/Waternet – company perspective

The study showed that it is certainly possible to build a 5MWth installation that converts WTP into 2.8MW electricity with an indirect gasifier and SOFC. However, the system is not feasible at the moment with an LCOE of 23.05 eurocent/kWh and an NPV of -38.20 Meuro. This could decrease to 15 eurocent/kWh with a net present value of -18.45 Meuro when the SOFC market grows to 100GW cumulative global shipments. In both cases the system does not reach break even, which makes it a bad investment from an economic perspective. Subsidy could help, but with a subsidy of 10.38 eurocent/kWh as analysed in this research, the NPV is still -2.3 million euros. Only when the economic analysis is solely based on OPEX cost, the system becomes profitable at an LCOE of 4.51 eurocent/kWh with an NPV of 2.41 million euro.

An alternative option could be to only dry the WTP to 25% moisture and then incinerate it at the AEB with an LCOE of 0.802 eurocent/kWh and an NPV of 6.35 million euro. In this case, the electricity yield will be 1.8 MW. Although the incineration option is more feasible, the electric efficiency of the gasifier-SOFC system is 57.2%, which makes it a promising option for high efficient electricity production (from waste) in the - near - future. Therefore, the recommendations based on the study are:

- Closely watch the SOFC price development, the technology is now in an early commercial phase and learning will lead to reduced system cost. However, learning will not reduce the cost enough to make the system feasible, so investigating more subsidy options is recommended as well.
- Investigate ways to increase the scale of the project by combining different waste (cellulose) streams or combining WTP from different locations. This will increase the feasibility of the project.
- When in a later stage a more detailed feasibility study is done, parameters that need most attention are:
  - o SOFC investment costs: The costs are currently based on open source data, mainly from fuel cell market reports. It would be more accurate to obtain direct quotations from SOFC suppliers at the required scale.
  - o Water content and amount WTP: These two parameters are interrelated. The 60% water content is estimation from Waternet, but not confirmed with a reasonable amount of experiments. By pressing the WTP, the water content at the gate of the AEB can be confined to 60%. Though, this could affect the amount of waste toilet paper available. It is therefore advised to investigate what will be the exact amount of waste toilet paper available at fixed moisture content.

6.2.2. Government/society perspective

The analysis done in this report shows that the system is very promising in terms of converting a waste stream -in this case waste toilet paper to electricity. The electrical efficiency is high (57.2%) compared to other options such as waste incineration (max. 30%) or integrated coal/biomass gasification (39%-45%). Moreover, the project can make citizens aware of the waste they create and at the same time show how their waste can still be a source of energy.

Though, at the moment the project is unfeasible in economic terms due to the small scale and relatively new technologies used. Therefore it is difficult for companies to invest. To support the development and feasibility for this kind of high efficiency systems, it is recommended to:

- Invest in research on SOFC systems in order to reduce the SOFC system costs.
- Invest in research on more flexible gasifier systems. This will help to increase the scale of the system because different waste streams can be combined which enhances the feasibility of the system.
- Create a category within the SDE+ scheme for (biomass) gasification systems in combination with fuel cells for electricity production. Currently, fuel cells are not mentioned at all, while this study shows they can contribute to highly efficient renewable electricity production.

6.2.3. Research perspective

This study has shown that it is possible to design a WTP to electricity system based on a gasifier and SOFC fuel cell. The system can be built on 5MWth scale with a high electrical efficiency of 57.2%. However, it also shows that it is problematic to make such small waste/biomass to electricity projects profitable, even more when they comprise relatively new technology such as a SOFC. Based on the above, the recommendations are:

- Improve the SOFC technology. A longer SOFC stack lifetime would decrease the amount of stack replacements and therefore the overall SOFC investments. One way to do this is looking for better catalysts that prevent carbon deposition on the anode.
- Perform gasification experiments with more different types of biomass/waste and combinations of streams to investigate how the scale could be increased. A good starting point could be to gasify the WTP and verify the results of this study, followed by combining other waste (cellulose) streams in the same gasifier to investigate the flexibility of the gasifier. When the gasifier system appears to be too sensitive to slight changes in the input, more research could be done to develop a gasifier that has increased fuel flexibility.

7. Literature

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Appendix A – Preliminary calculation

A first calculation was conducted to get an idea of how many households can be powered by waste toilet paper. The starting point is a production of 25.000 tonnes of waste toilet paper (WTP) per year with 60% water content or 10.000 tonnes dry material (C. Reijken from Waternet, personal communication; Ruiken et al., 2010) from the WWTP in Amsterdam-West. The LHV and HHV of WTP are determined by ECN and given in Ruiken et al (2010) as 16.5 MJ/kg LHV and 17.85 MJ/kg HHV dry material. For this calculation, the LHV value was used. So the total energy in the WTP per year is 165 TJ. When the installation would work year-round, this means a scale of:

165 TJ $* 10^6$/(8760*3600) = 5.23 MW.

With a cold gas efficiency (how much of the energy in the feed is converted into product gas) of 80% for the gasifier (Fryda et al., 2008; van der Meijden, 2010), 4.18 MW of energy is left after gasification. Then for the fuel cell an electrical efficiency of 55% (Lower Heating Value) is taken from James et al. (2012). This gives an electrical output of 2.3 MWe or 20.17 GWh/y.

The electricity consumption of an average Dutch household is 3150 kWh/year in 2013 (CBS, 2015). This means that 6402 households can be provided with electricity from waste toilet paper.

Appendix B – Waste water treatment plant

Waste water from the sewage system and house hold waste water are collected and treated in waste water treatment plants (WWTP) before draining to surface water. In a general WWTP the wastewater follows a number of cleaning steps, which are explained below and visualized in Figure 17 (Ruiken et al. 2010). The difference with the new process to extract the WTP is a change in the second step.

There are two options; the first option is to completely or partly replace the primary sedimentation tanks with a sieve or filter that mainly filters out waste toilet paper. This alternative process is presented in Figure 18, were step 2A is replaced by step 2B. As most WWTPs already have the primary sedimentation tanks, another option is to extract WTP from the primary sludge, and step 2A is followed by step 2C (Figure 19).

Waste water treatment steps

1. Bar screen
   To remove all large objects as leaves, cans, plastic, sanitary pads, etc. This garbage is qualified as hazardous garbage and is pressed to remove some water and then it is processed by a waste treatment or incineration company.

2. A) Primary sedimentation tanks
   The water enters the cylindrical tank from the middle under the surface and is circulated in order to let heavy material (such as sand) sink. Fats will float at the top. Both the primary sludge from the bottom and the fats are removed. Fats can sometimes be recycled. The primary sludge could then be processed in a fermentation reactor to produce biogas. Remaining sludge is dehydrated and afterwards transported to a waste processing company to be burned.

   B) Sieves/Filters
   Instead of the primary sedimentation tanks, big filters with a mesh size of < 0.5 mm can be installed. In this case, mostly toilet paper, hair and sand are filtered out before entering aeration tanks. The product is called 'zeefgoed' but in this report called WTP. This method could work for newly build WWTPs.

   C) Primary Sludge extraction
   When primary sedimentation tanks are already in place, 'zeefgoed' could be extracted after primary sedimentation because it is not economically viable to install sieves in that case. Waternet is currently testing this method and a report will follow at the end of 2015 (C. Reijken, personal communication).

3. Aeration tank
   The next step is to clean the water by using bacteria that are already present. The bacteria use the remaining biological waste as nutrients. In this tank the water is circulated faster than in the primary sedimentation tanks and with additional airflow in order for the bacteria to receive enough oxygen. The temperature is between 10-22°C and the water stays in this tank for one day.

4. Secondary sedimentation tanks
   The bacteria have become heavier after the former step and will sink to the bottom of this cylindrical tank. The bacteria are called 'active sludge' and part of it is transported back to the aeration tank. The surplus of active sludge is treated in the same way as the primary sludge. The water is now considered clean and in most cases this is the last cleaning step.

5. Sand filtration (optional – not included in figures)
   Some WWTP do consider it necessary to perform sand filtration after the sedimentation tanks. Sand filtration can be included as an extra step to filter out remaining small particles. If this step is necessary will depend on the requirements for the fresh water quality in a specific area and on the kind of water body that the effluent is discarded to.

6. Discharge of the water (effluent)
   The cleaned water is called effluent and finally flows into a body of surface water. The water quality of the effluent is checked regularly.
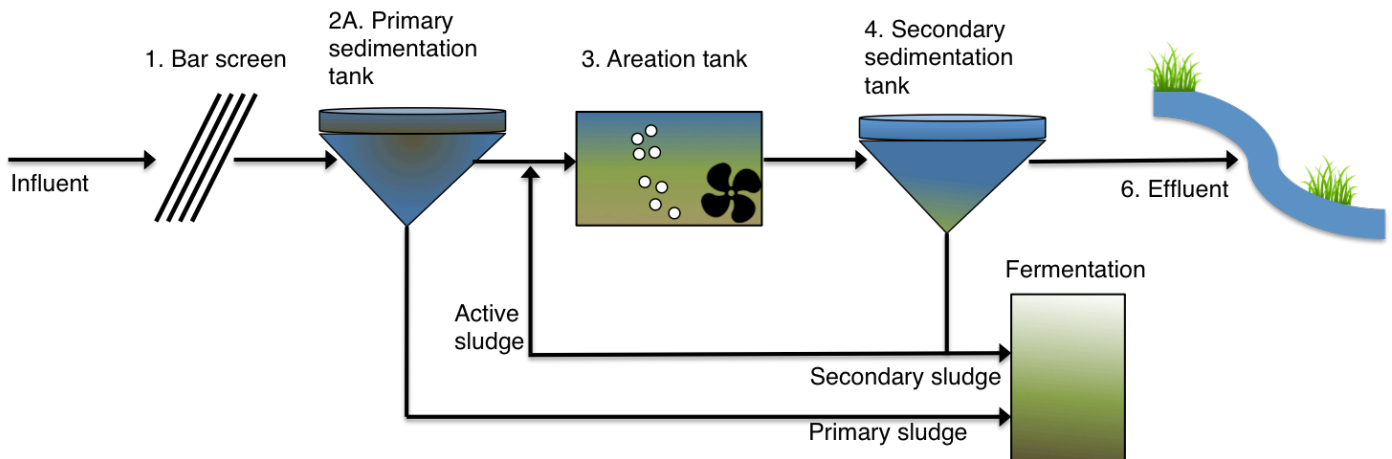
Figure 17 Visualisation of the processing steps in a Waste Water Treatment Plant. Based on Ruiken et al (2010).
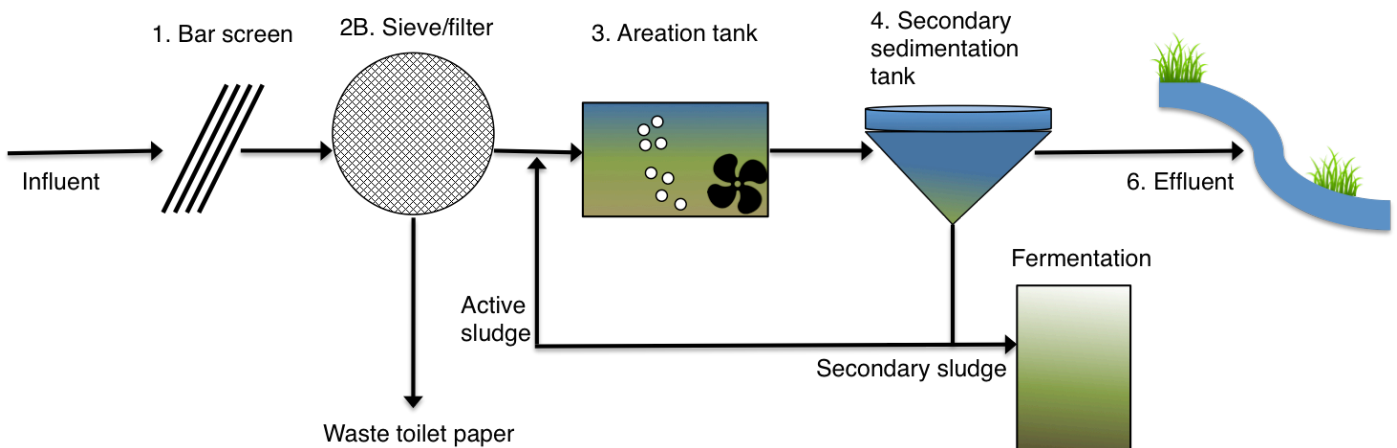
Figure 18 Visualistation of the processing steps in a Waste Water Treatment Plant with sieve for waste toilet paper. Based on Ruiken et al. (2010).

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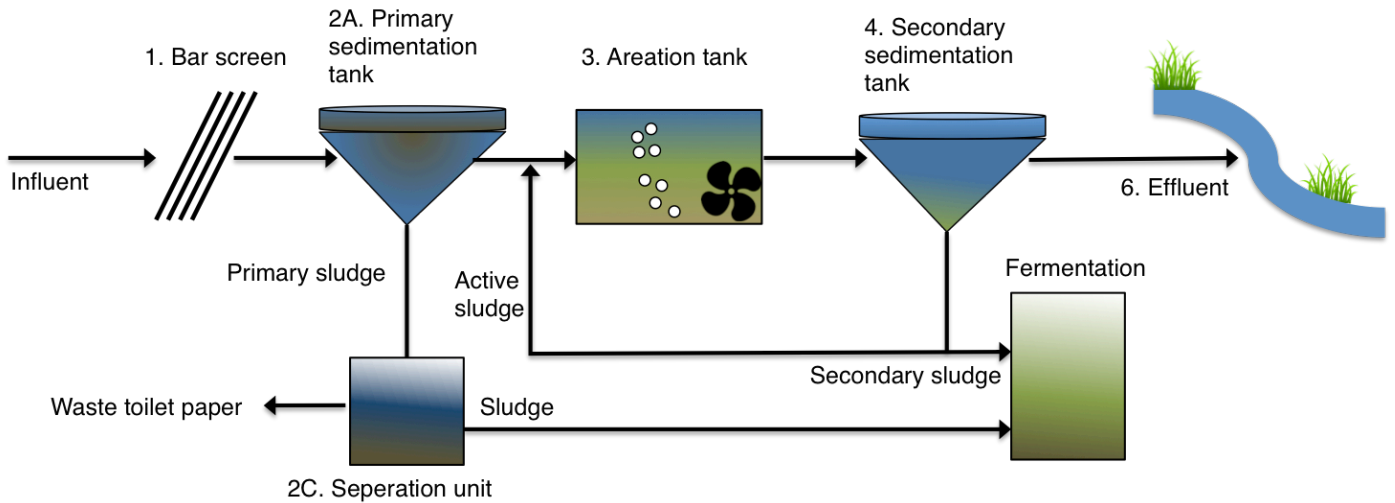
Figure 19 Visualization of the processing steps in a Waste Water Treatment Plant with separation unit for waste toilet paper. Based on Ruiken et al. (2010).

The reason for separating the WTP from the waste water before it enters the next steps of the treatment process is to improve the WWTP. According to the STOWA publication about zeefgoed (Ruiken et al. 2010), some advantages could be:

- It is expected that the amount of failures will be less.
  Normally, a combination of fibers, hair and cloth that did pass the bar screen form balls of material that sometimes block the water- or sludge stream. By early separation, these problems are solved.
- The cleaning process becomes more energy efficient.
  Not all cellulose that is present in the water breaks down during the cleaning steps. The waste-sludge still has 15-23% cellulose content, which is just extra waste. Filtering out the cellulose in an earlier stage can thus reduce the amount of waste sludge. Moreover, the aeration process could be accelerated when the bacteria do not have to break down cellulose anymore.
- Waste processing costs decrease.
  Costs for dehydrating and processing of waste sludge are about €450/ tonne dry matter. Compared to the estimated costs for processing of zeefgoed of about € 20-100 / tonne dry matter, this difference could make it more profitable to use sieves/filters.
- For this research, the WWTP at Amsterdam-West (next to the AEB) is considered and here the new method (2C) from will be the most appropriate

Appendix C – Overview of Fuel Cell Suppliers

See Table 24 for an overview of (almost) commercial SOFC suppliers. The development of SOFC fuel cells lay a bit behind the other type of high temperature fuel cells - Molten Carbonate Fuel Cells - due to their shorter lifetimes. Now the lifetime is increasing above 30.000 hours, commercialization of the SOFC fuel cell is slowly starting. For this research, only fuel cell stacks from at least 50kW meant for stationary applications are selected from (Mcphail et al., 2012, 2013).

Table 24 Overview of (almost) commercial SOFCss and some characteristics

	Bloom Energy[a]	LGFCS[b]	Fuel Cell Energy[c]	Mitsubishi Heavy Industries[d]	Convion/ Wärtsilä[e]
Operating T	nd	nd	700-1000	900-1000	nd
Type	planar	planar	planar	cylindrical	planar
Feedstock	Natural gas/biogas	Natural gas	Natural gas	Natural gas	Natural gas, city gas, biogas
State of commercialization	Commercial Sells to i.e. Ebay	Quite unclear. No website available	Commercial	Demonstration, not really commercialized	Demonstration, moving to commercialization in 2015-2016
Size of stacks	250 kW	1 MW (4x 250kW)	300 kW, 1.4 MW or 2.8 MW	250 kW	50-300 kW
Expected lifetime	10 years (warranty)	Should be >40.000 hours	nd	Until now 4000 running hours	nd
Electrical efficiency	52-60%	>55%	47%	52% (LHV)	53% (LHV) with 53 kW
Total efficiency	nd	nd	nd	65% (HHV, combined with gas turbine)	80% (exhaust gas at 40 degrees)
Heat available for recovery	nd		500 MJ/h – to 120 degrees 850 MJ/h – to 50 degrees	nd	Exhaust gas T 222 degrees
Data on costs	Yes (100kw 700.000 – 800.000 dollar)	nd	Not readily available, but is possible	No	Not yet
Country	US, California	Unclear	US, Connecticut	Japan	Finland

[a](Basset, England, Li, Weinberger, & Wong, 2010; Bloom Energy, 2015) [b](Goettler, 2014; Mcphail et al., 2013)
[c](Fuel Cell Energy, 2013; Ghezel, n.d.) [d](Mcphail et al., 2013; Mitsubishi Heavy Industries, 2015) [e](Convion, 2015)

Appendix D – Interview summaries

Notes from Interview with Chris Ruijken –Waternet, 19 October 2015
Goal: Learning more about the 'production' of waste toilet paper, the exact amounts that are available and the elemental analysis.

Most important outcomes:
- At the moment, there are no sieve-installations working. The WWTP in Blaricum has stopped working and will hopefully start working again soon.
- Instead of sieving, the cellulose waste will be extracted from the primary slugde. This will make it more dirty and thus not as clean as 'zeefgoed', but for now I can assume the contaminants and composition as know for zeefgoed.
- There was no elemental analysis done, just the contaminant analysis. Chris gave the following numbers as an indication from the dry matter: 70% cellulose, 8-10% ash, 4-5% fat/protein, 6% water. Rest is unknown.
- 10.000 ton/y (dry, about 6% moisture) of WTP is possible. Currently there is no installation in place, but for the research I may assume that this is the production from the WWTP of Amsterdam-West.

Notes from Interview with Berend Vreugdenhil – ECN, 23 November 2015
Goal: Discussing what would be the best type of gasifier to convert toilet paper. Asking questions about modelling of gasification, catalysts, contaminants in the product gas and cleaning equipment.

Most important outcomes:
- An indirect gasifier is the best option in for coupling with a fuel cell, since a very high circulation rate of fuel gas in the fuel cell would be needed when there is 80% nitrogen in the gas. The other option would be to install an air separation unit, but from an economic perspective this is no option at all at this scale. The Milena gasifier is thus a good option for this process.
- The material has to be dried to a maximum of 25% moisture for the Milena gasifier.
- Modelling the gasifier in Aspen makes no sense if you are not able to verify the model with experiments or adapt the model to literature. The equilibrium blocks will give results that are nonsense and by pre-setting the concentrations for various components (such as hydrocarbons) you can as well just use literature data.
- The process chain could look like this: Drying – Milena – Cyclone – OLGA tar removal – HDS (for thiophenes) – Sorbent material – ZnO – Prereformer (for higher hydrocarbons) – SOFC
- Phosphorous elements are probably bounded to ash in the form of CalciumPhosphate
- No catalysts are needed in the gasifier, for toilet paper sand could work as bed material (and also olivine, but that is more expensive) and 850°C operating temperature should work as well.
- Berend thinks that a better business case for the AEB would be: MILENA gasification -> OLGA gas cleaning -> boiler -> district heating system

Notes from the Interview/discussion with Bob van der Zwaan – ECN/UvA, 06-01-2016.
Goal: Discussing my project, with a focus on discussion of the economic analysis. Validate the method I proposed for the economic analysis.

Most important points from the meeting with Bob van der Zwaan:
- It doesn't matter if you first scale the system and then convert to euros or the other way around, as long as you write down what you did.
- LCOE is the most important number in the analysis.

- Better only look at your commercial system, and do a good analysis on future costs by means of scaling instead of looking at a 'research scenario'.
- Do a sensitivity analysis on the assumptions you make, so for example the cost escalation factor, the lifetime of the system and the learning rate you choose.

Notes from the Interview/meeting with Martin 't Hoff, 08-01-2016
Goals: Obtaining cost data for MILENA and OLGA

Most important points from the meeting:
- We discussed the system design and made some final changes:
  o The HDS reactor has to work on 3 bars, so a compressor is added.
  o Because a compressor would lead to water condensation at the current water content in the gas, water will be removed first in a water scrubber. Water content will be reduced to 6%.
  o Because the water scrubber removes HCl, the $NaHCO_3$ reactor can be removed from the design.
  o One ZnO reactor is enough for H_2S removal at this scale, if it includes an extra guard bed to prevent break through.
- Removal rate of NH_3 in water scrubber: ca 40%, HCl removal rate: 98%.
- Dryer air can be at 120°C
- Cost specifications of the whole system at the right scale from dryer until ZnO- reactor were obtained, see section 3.6.1.
- An estimation of the O&M (fixed and variable) costs was obtained, see section 3.6.3.

Appendix E – Data used in model

Table 25 LHV/HHV of WTP dried, with 60% moisture and 25% moisture

	dried 95degrees (wt%)	After pressing (60% moisture)	After dryer (25% moisture)
HHV (MJ/kg)	17.49	7.00	13.12
LHV (MJ/kg)	16.13	4.99	11.49

Formulas for conversion of HHV/LHV dry based to wet based (Nieuwlaar et al., 2014):

$$E_{HHV,wb} = E_{HHV,db} \cdot (1 - mc)$$

$$E_{LHV,wb} = E_{LHV,db} \cdot (1 - mc) - E_{w,evap} \cdot mc$$

Where wb=wet based and db=dry based, mc = moisture content (mass fraction)

Table 26 LHV and HHV values of product gas components

Heating values	HHV (kJ/kg)	LHV (kJ/kg)
CO	10100[1]	10100[1]
H2	141800[1]	120000[1]
CO2	-	-
O2	-	-
H2O	-	-
CH4	55530[1]	50050[1]
N2	-	-
Ar	-	-
C2H2	49970[1]	48280[1]
C2H4	54156[2]	51019
C2H6	51900[1]	47520[1]
C3H6	48885[2]	45749
C3H8	50324[2]	46334
C4H10	49150[1]	45370[1]
C6H6	41800[1]	40100[1]
C7H8	42400[1]	40500[1]
H2S from fuel	16514	15224[3]
COS		
NH3 from fuel	22484	18610[3]
HCl from fuel		
HCN from fuel	24586	23772[3]
thiophene		

Sources: [1](Cengel & Boles, 2014) [2](Verkerk et al., 2008) [3](Nao, n.d.)

Table 27 Air compostion

Air	vol/mol %
N_2	77.30%
O_3	20.74%
Ar	0.92%
CO_2	0.03%
H_2O	1.01%
mass (g/mol)	28.86
kg/mol	0.03

Table 28 Tar & char composition, taken from van der Meijden (2010).

		Char	Tar
C	wt% daf	91%	93%
H	wt% daf	1%	6%
O	wt% daf	8%	1%
N		30% of N in fuel	
HHV	MJ/kg	32.33[1]	39.40
LHV	MJ/kg	32.10[1]	38.10

[1]Derived from the excel model

Table 29 Specific heat capacity (kJ/kg/K) of solids and gasses used for energy balance calculations, plus heat for water evaporation.

	Value (kJ/kg/K)	Derived at temperature
c_{water}	4.18	
$c_{p, WTP}$[1]	1.90	15°C
$c_{, ash + char}$[1]	1.01	850°C
c_{tar}[1]	9.04	800°C
$c_{raw product gas}$[2]	2.189	850°C
$c_{raw product gas}$[2]	2.159	800°C
$c_{raw product gas}$[2]	1.878	400°C
$c_{raw product gas}$[2]	2.189	80°C
$c_{water scrubber gas}$[2]	1.464	50°C
$c_{water scrubber gas}$[2]	1.566	145°C
$c_{water scrubber gas}$[2]	1.759	350°C
$c_{water scrubber gas}$[2]	1.802	400°C
$c_{water scrubber gas}$[2]	1.921	550°C
$c_{water scrubber gas}$[2]	2.085	800°C
$c_{combustor flue gas}$[2]	1.206	900°C
$c_{combustor flue gas}$[2]	1.001	30°C
$c_{p,air}$[3]	1.004	0°C
$c_{p,air}$[3]	1.006	15°C
$c_{p,air}$[3]	1.022	120°C
$c_{p,air}$[3]	1.059	320°C
$E_{w,evap}$ **(water evaporation)**	2.442 **MJ/kg**	25°C

[1] The specific heat capacity of WTP is not known, and therefore derived from the model of van der Meijden (2010) as if it was wood pellet. For char + ash and tar, the same assumption was made. Also, the c was assumed to be constant with varying temperature
[2] Derived from Aspen as explained in section 32
[3] Derived from (Cengel & Boles, 2014)

Table 30 Conversion rate of contaminants in different parts of the cleaning equipment

Equipment / *compound*	Removal rate	Part of / *converted to*
Water scrubber		
Water[1]	6 vol% after scrubbing	**Waste water**
HCl[1]	98%	**Waste water**
NH$_3$[1]	40%	**Waste water**
HDS		
COS[2]	100%	*CO + H$_2$S +*
Thiophene (C$_4$H$_4$S)[2]	100%	*C$_4$H$_{10}$ + H$_2$S*
HCN[2]	98%	*CH$_4$ + NH$_3$*
C$_2$H$_2$[2]	100%	*C$_2$H$_6$*
C$_2$H$_4$[2]	100%	*C$_2$H$_6$*
C$_3$H$_6$[2]	100%	*C$_3$H$_8$*
ZnO		
H$_2$S[3]	Reduced to 0.1 ppmv (removal rate of 99.999%)	*ZnS*

[1] **Martin van 't Hoff (Personal communication). Water was validated with van der Meijden who reported 7% water in the gas after scrubbing. For HCl van der Meijden (2010) only notes that 'most HCl is removed and part of the NH$_3$' (par. 6.5.5).**
[2] **(Rabou & Almansa Aranda, 2015)**
[3] **ZnO is known for its good quality H$_2$S removal < 1ppm as stated by (van Paasen et al., 2006; Zwart, 2009). (Rabou & Almansa Aranda, 2015) even reported that all S compounds were reduced to <0.1 ppm. In this report, the H$_2$S concentration was chosen to be reduced to 0.1ppmv.**

Appendix F – Mass & Energy balances

Important notes:
- For all energy balances: the reference temperature is 0°C
- Inlet T for air/WTP from outside is 15°C

Dryer

The purpose of the dryer was to decrease the water content in the WTP from 60% to 25%. The most important number that had to be obtained was the amount of energy needed to evaporate the water and heat up both water, biomass and air.

The mass balance (Table 31) only contains the biomass in- and out- stream and water in air out stream, the amount of air needed was not calculated. The mass balance was solved by using the mass balance for water and for the total streams.

Table 31 Mass balance dryer

Mass balance			
IN	**kg/hr**	**OUT**	**kg/hr**
WTP	3567.35	Dried WTP	1902.59
Air		Air	
		Water in air	1664.76
Total	3567.35		3567.35

Table 32 Energy balance dryer, heat supplied by air is calculated by difference.

Energy balance							
IN	**kg/s**	**LHV (kW)**	**HHV (kW)**	**OUT**	**kg/s**	**LHV (kW)**	**HHV (kW)**
WTP	0.99	4942.53	6932.15	Dried WTP	0.53	6071.80	6932.15
Qsens WTP	0.99	28.27	28.27	Qsens WTP	0.53	15.08	15.08
Heat supplied by air[1]	-	2409.64	2409.64	Q condensed water[2]			1129.26
				Q Water in Air[3]	0.46	1293.57	1293.57
Total		7380.44	9370.07			7380.44	9370.07

[1] The *heat supplied by air* was calculated by difference to solve the energy balance
[2] The heat from condensed water (*Q condensed water*) is the difference between the HHV and LHV of the wet and dried WTP.
[3] The *Q water in air* contains the energy needed to heat up the water in the biomass from 15°C to 100°C plus the energy needed to evaporate this water, see Table 33 for constants used.

Table 33 Temperatures used for energy balance dryer

T WTP in	15	°C
T WTP out	15	°C
T steam out	100	°C
T air in	120	°C

Gasifier

Table 34 Mass balance gasifier

Mass balance			
IN	kg/hr	**OUT - to settling chamber**	kg/hr
WTP dried	1902.59	Raw producer gas	1887.41
CO2	221.29	Tar	39.87
		Char	125.63
Recycle from combustor	32.17	Ash	67.78
		OUT - to combustor	
		Recyling to combustor[1]	15.54
Total	**2156.05**		**2136.22**

[1]Only gas + tar

Table 35 Energy balance gasifier

Energy balance							
IN	kg/s	LHV (kW)	HHV (kW)	**OUT**	kg/s	LHV (kW)	HHV (kW)
Qsens WTP dried	0.53	15.08	15.08	Qsens producer gas	0.52	918.04	918.04
WTP	0.53	6071.80	6932.15	Q cond. Water			391.29
Q from combustor (indirect)	-	607.14	610.74	Producer gas	0.52	5404.36	5851.03
				C loss to char	0.03	1120.32	1120.32
				Tar loss	0.01	421.91	421.91
				Sensible heat as + char	0.03	28.22	28.22
				Heat loss[1]		0.00	0.00
Total[2]		6694.01	7557.97			7892.85	8753.21

[1] It is assumed that there is no heat loss in the gasifier, but there is heat loss in the combustor.
[2] The difference is 1195 kW between the in and outflow. Probably this is caused by using incorrect numbers for the specific heat capacity of tar, char and ash. Those numbers are derived from van der Meijden (2010) as if wood is gasified. Also, the cp for sensible heat does not change with temperature.

Table 36 Temperatures used for energy balance gasifier

T$_{WTP}$ in	15	°C
T gas/tar/char/ash out	800	°C

Settling chamber gasifier

In the settling chamber within the MILENA gasifier, 90% of the char & ash are settled and send to the combustor. No energy balance is given here because nothing happens with the energy streams.

Mass balance			
IN	**kg/hr**	**OUT - to HX2**	**kg/hr**
Producer gas	1887.41	Producer gas	1887.41
Ash	67.78	Ash	6.78
Char	125.63	Char	12.56
Tar	39.87	Tar	39.87
		OUT - to combustor	
		Ash	61.00
		char	113.06
Total	2120.68		2120.68

HX – 2

For heat exchangers, no mass balance is given because nothing changes in the mass flows. In the energy balance, only the sensible heat (Qsens) is given as the energy content in the components stay constant. The gas is cooled from 800°C to 400°C.

Energy balance							
IN	**kg/s**	**LHV (kW)**	**HHV (kW)**	**OUT – to cyclone**	**kg/s**	**LHV (kW)**	**HHV (kW)**
Qsen Producer gas	0.53	905.50	905.50	Qsen Producer gas	0.53	393.81	393.81
Qsen ash	0.00	1.52	1.52	Qsen ash	0.00	0.76	0.76
Qsen char	0.00	2.82	2.82	Qsen char	0.00	1.41	1.41
Qsen tar[1]	0.01	?	?	Qsen tar	0.01	?	?
				Heat	?	**513.86**	**513.86**
Total		917.15	917.15			909.84	909.84

[1] The specific heat capacity (kJ/kg/K) could not be derived from the model of van der Meijden (2010) so is left out of the energy balance.

Cyclone

In the cyclone 90% of the ash and char are removed from the gas, heat loss is assumed to be negligible. Therefore only the mass balance is given.

Table 37 Mass balance cyclone

Mass balance			
IN	**kg/hr**	**OUT - to OLGA**	**kg/hr**
Producer gas	1887.41	Producer gas	1887.41
Ash	6.78	Ash	0.68
Char	12.56	Char	1.26
Tar	39.87	Tar	39.87
		OUT - to combustor	
		Ash	6.10
		Char	11.31
Total	1946.62		1946.62

Combustor

In the combustor the char and tar are completely burned. Air for burning comes via OLGA together with the light tars. Heavy tars come via a separate stream from OLGA. Char and ash come from the settling chamber within the MILENA gasifier as well as via the cyclone. 1% of the flue gas is recycled to the gasifier and 0.8% of the product gas enters the combustor.

Table 38 Mass balance combustor

Mass balance			
IN	**kg/hr**	**OUT**	**kg/hr**
Light tar	26.58	Flue gas	4547.01
Heavy tar	13.29	White ash	67.10
Ash - settling chamber	61.00		
Char - settling chamber	113.06		
Ash cyclone	6.10	Flue gas to gasifier	45.93
Char cyclone	11.31		
Air – via OLGA	4413		
Product gas from gasifier	15.22		
Total	**4660.04**		**4660.04**

Table 39 Energy balance combustor

Energy balance							
IN	**kg/s**	**LHV (kW)**	**HHV (kW)**	**OUT**	**kg/hr**	**LHV (kW)**	**HHV (kW)**
Sensible heat char + ash - settling chamber	0.05	39.10	39.10	Sensible heat flue gas	1.29	1391.32	1391.32
Qsens char + ash - cyclone	0.01	1.96	1.96	Heat loss[2]		98.85	98.85
Qsens tar – air[1]	0.007	?	?	Q condensed water			22.32
Qsens air with tar	1.23	498.59	498.59				
Char	0.035	1109.12	1117.04	Heat supply to gasifier[3]		607.14	610.74
Tar - air	0.01	281.28	290.87	Qsens ash	0.0188	16.96	16.96
Tar - oil	0.004	140.64	145.44				
Producer gas from gasifier	0.004	43.58	47.19				
Total		**2114.27**	**2140.19**			**2114.27**	**2140.19**

[1] The specific heat capacity (kJ/kg/K) could not be derived from the model of van der Meijden (2010) so is left out of the energy balance.
[2] Heat loss is assumed to be 2% of the WTP input (LHV) (van der Meijden, 2010)
[3] The heat supply to the gasifier is detrimed by difference to solve the energy balance

Table 40 Temperatures used for combustor energy balance

T char+ash – settling chamber	800	°C
T char+ash cyclone	400	°C
T light tar/air	380	°C
T producer gas	800	°C
T flue gas	900	°C
T ash	900	°C

HX3

In this heat exchanger the flue gas from the combustor is cooled down from 900°C to 35°C.

Energy balance							
IN	kg/s	LHV (kW)	HHV (kW)	OUT - to HDS	kg/s	LHV (kW)	HHV (kW)
Qsens flue gas	1.28	1384.93	1384.93	Qsens flue gas	1.28	44.77	44.77
Qsens ash	0.02	16.02	16.02	Qsens ash	0.02	0.66	0.66
				Heat		1355.52	1355.52
Total		1400.95	1400.95			45.42	45.42

HX4

This heat exchanger heats up air for the OLGA system from 15°C to 380°C.

Table 41 Energy balance HX4

Energy balance							
IN	kg/s	LHV (kW)	HHV (kW)	OUT - to OLGA	kg/s	LHV (kW)	HHV (kW)
Sensible heat gas	1.23	18.46	18.46	Sensible heat gas	1.23	498.59	498.59
Heat needed		480.14	480.14				
Total	1.23	18.46	18.46		1.23	498.59	498.59

OLGA

In the OLGA tar removal system, both light tars are removed and travel together with an air stream to the combustor while heavy tars are removed with scrubbing oil, see also 2.3.2.

Table 42 Mass balance OLGA

Mass balance			
IN	kg/hr	OUT - to Water scrubber	kg/hr
Producer gas	1887.41	Producer gas	1887.41
Tar	39.87		
Air	4413.49	**OUT - to combustor**	
Stripper oil bleed[1]	0.34	Heavy tar - oil	13.29
		Light tar - air	26.58
		Oil	0.34
		Air	4413.49
Total	6341.10		6341.10

[1]There is a small bleed of oil. In the mass and energy balance of the combustor this oil bleed is not taken into account.

Table 43 Energy balance OLGA

Energy balance							
IN	kg/s	LHV (kW)	HHV (kW)	OUT - to water scrubber	kg/s	LHV (kW)	HHV (kW)
Qsens gas	0.53	416.23	416.23	Sensible heat gas		104.06	104.06
Qsens tar	0.01			Producer gas[1]	0.53	5404.36	5851.03
Qsens air	1.23	498.59	498.59				
Producer gas	0.53	5404.36	5851.03	**OUT - to combustor**			
Tar	0.01	421.91	436.31	Qsens air[2]	1.23	498.59	498.59
				Qsens light tar			
				Qsens heavy tar			
				Light tar	0.01	281.28	290.87
				Heavy tar	0.00	140.64	145.44
				Heat loss[3]		312.17	312.17
Total		6741.10	7202.16			6741.10	7202.16

[1] The heating value of the producer gas does not change when tar is removed, because tar was not included in the LHV/HHV of the producer gas
[2] Is it assumed here that the air is not cooled during stripping and goes to the combustor still at 380°C.
[3] In OLGA the gas is cooled down from 400°C to 80°C, which will result in a heat loss.

Water scrubber

In the water scrubber, the water content is reduced and HCl and NH_3 are removed. The gas is also cooled from 80°C to 50°C.

Mass balance			
IN	**kg/hr**	**OUT - to Compressor/HX5**	**kg/hr**
Producer gas	1887.41	Cleaned producer gas	1357.59
Water in producer gas	*588.94*	*Water*	*63.58*
HCl in producer gas	*2.60*	*HCl*	*0.05*
NH3 in producer gas	*4.76*	*NH3*	*2.85*
		Waste water stream	
		Water removed	*525.37*
		HCl removed	*2.55*
		NH3 removed	*1.90*
Total	**1887.41**		**1887.41**

Table 44 Energy balance water scrubber

Energy balance							
IN	**kg/s**	**LHV (kW)**	**HHV (kW)**	**OUT - to compressor/HX5**	**kg/s**	**LHV (kW)**	**HHV (kW)**
Sensible heat gas	0.53	67.91	67.91	Sensible heat gas	0.38	27.60	27.60
Producer gas	0.53	5404.36	5851.03	Producer gas	0.38	5377.81	5821.07
				OUT (lost)			
				Sensible heat waste water	0.15	30.50	30.50
				NH3 out	0.00	12.56	11.88
				Heat loss[1]		23.79	27.89
Total		5472.27	5918.94			5472.27	5918.94

[1] Because the gas is cooled down from 80°C to 50°C, some heat is lost.

Compressor

In the compressor the pressure of the producer gas rises from 1 bar to 3 bar and therefore also temperature rises from 50°C to 145°C (Royal Dalhman, personal communication). The work of the compressor is calculated by thermodynamic calculations with as assumptions an isentropic efficiency of 80% and a mechanical efficiency of 98% taken from van der Meijden (2010). The mass balance does not change.

Table 45 Energy balance compressor

Energy balance							
IN	kg/s	LHV (kW)	HHV (kW)	OUT - to compressor/HX 5	kg/s	LHV (kW)	HHV (kW)
Qsens gas	0.38	22.08	22.08	Qsens gas	0.38	85.63	85.63
Producer gas	0.38	5377.81	5821.07	Producer gas	0.38	5377.81	5821.07
Electricity		94.47	94.47	Heat loss		30.93	30.93
Total		5494.37	5937.62			5494.37	5937.62

HX5

In this heat exchanger the temperature of the gas has to rise from 145°C to 350°C.

Energy balance							
IN	kg/s	LHV (kW)	HHV (kW)	OUT - to HDS	kg/s	LHV (kW)	HHV (kW)
Qsens gas	0.38	85.63	85.63	Qsens gas	0.38	232.22	232.22
Producer gas	0.38	5377.81	5821.07	Producer gas	0.38	5377.81	5821.07
Heat		146.60	146.60				
Total		5610.04	6053.29			5610.04	6053.29

HDS

In the HDS reactor thiophene, COS, HCN and hydrocarbons are converted, see for more information 2.3.4. Conversion rates can be found in Table 30.

Mass balance			
IN	kg/hr	OUT - to HX6/ZnO	kg/hr
Producer gas	1357.59	Cleaned producer gas	1372.90
Thiophene in producer gas	0.68	*Tiophene*	*0.00*
COS in producer gas	0.48	*COS*	*0.00*
C2H4 in producer gas	64.34	*C2H4*	*0.00*
H2 in gas	30.29	*C2H2*	0.00
HCN in gas	1.52	*C3H6*	0.00
C2H2 in producer gas	4.52	HCN	0.03
C3H6 in producer gas	2.34	C4H10	0.47
C2H6 in producer gas	4.60	CO	0.23
CH4 in gas	111.50	H2S	0.55
		C2H6 total	78.79
		C3H8	2.45
		CH4 total	112.43
		H2 total in gas	*24.42*
		NH3	0.99
Total	220.27		220.35

In the HDS, the temperature rises because of the hydrogenation of hydrocarbons. Because the c_p of the producer gas was known (see appendix E) as well as the energy content of the gas after treatment in the HDS (see method section), the temperature of the outgoing stream could be determined with Equation 6 and was found to be 553°C.

Energy balance							
IN	kg/s	LHV (kW)	HHV (kW)	OUT - HX6/ZnO	kg/s	LHV (kW)	HHV (kW)
Qsens gas	0.38	232.22	232.22	Qsens gas	0.38	400.57	400.57
Producer gas	0.38	5377.81	5821.07	Cleaned gas - after HDS	0.38	5209.47	5652.75
Total		5610.04	6053.29			5610.04	6053.32

HX6

In this heat exchanger the gas is cooled from 553°C to 350°C before entering the ZnO reactor.

Energy balance							
IN	kg/s	LHV (kW)	HHV (kW)	OUT - ZnO	kg/s	LHV (kW)	HHV (kW)
Sensible heat gas	0.38	5610.04	6053.32	Sensible heat gas	0.38	234.84	234.84
Producer gas after HDS	0.38	5610.04	6053.32	Producer gas after HDS	0.38	5209.47	5652.75
				Heat		170.24	170.24
Total		5444.31	5887.59			5444.31	5887.59

ZnO

In the ZnO reactor H_2S is converted to ZnS. The temperature stays constant at 350°C.

Table 46 Mass balance ZnO

Mass balance			
IN	**kg/hr**	**OUT - to HX7 /SOFC**	**kg/hr**
Producer gas	1372.90	Cleaned product gas	1370.16
H2S in producer gas	5.80	*H2S in producer gas*	*0.00*
		H2O added	*3.06*
ZnO used	13.84	Spend ZnS	*16.58*
Total	**1386.74**		**1386.74**

Table 47 Energy balance ZnO reactor

Energy balance							
IN	**kg/s**	**LHV (kW)**	**HHV (kW)**	**OUT - SOFC**	**kg/s**	**LHV (kW)**	**HHV (kW)**
Sensible heat gas	0.38	234.84	234.84	Sensible heat gas	0.38	234.84	234.84
Cleaned gas after HDS	0.38	5268.20	5716.48	Cleaned gas - after ZnO	0.38	5268.20	5716.48
				Energy loss to ZnS[1]		14.06	15.25
Total		**5503.04**	**5951.32**			**5503.04**	**5951.32**

[1]Because the H_2S is converted its energy is not part of the product gas anymore

SOFC

In the fuel cell the cleaned gas is converted into heat and electricity. Electrical efficiency is 55% LHV and efficiency to heat 31%, see 4.2.4 for more information.

Table 48 Energy balance SOFC.

IN	kg/s	LHV (kW)	OUT	kg/s	LHV (kW)
Cleaned gas after ZnO	0.38	5254.14	Electrity	0.38	2889.78
Qsens gas	0.38	234.84	Heat	0.38	1628.78
			Heat loss		**970.42**
Total		5488.98			5488.98

Appendix G – Fuel cell Investment cost

The fuel cell investment cost were gathered from different sources and are summarized in able 50, including the system parts that were said to be part of the cost estimate. As can be seen, the costs vary highly among the different sources. Prices were derived according to the method as described in 3.6.1. The dollar-euro exchange rates used (Table 49) were calculated via (OANDA, 2016), the average exchange rate over a whole year was taken.

Table 49 Dollar - Euro exchange rates

	2001	2003	2007	2012	2013	2014	2015
Dollar - Euro	1.1165	0.8846	0.7306	0.778	0.7531	0.7538	0.9142

For the calculations we decided to use the 1.4 MWe/1.1MWth CHP system from Ammerman et al. (2015) for four reasons. Firstly, the publication is very recent and detailed which makes it quite reliable. Secondly it is based on the European fuel cell market instead of for example the US market, which makes it more applicable for the Dutch business case. Thirdly, the publication was made in collaboration with fuel cell companies in Europe (Abengoa, AF FC Energy, Ballard Power Syste ms, Baxi Innotech, Ceramic Fuel Cells CFCL, Ce res Power, Convion Oy, elcore, Dantherm Power, FuelCell Energy Solutions, Hydrogenics, IE- CHP, IRD, Proton Motor, SOFC Power, Sunfire, Topsoe Fuel Cell, Vaillant, Viessmannn) who also delivered cost data.

able 50 SOFC system costs

	Invest ment cost (M€)	Instum entatio	Piping	AC/DC inverter	Turbine	BOP	Pre-reforme	Water-system	Heaters	Sensors
Electricity only (1MW)[1]	12.43	x	x	x	x	x	x	x	x	x
CHP (1.4 MWe/1.1 MWth)[1]	11.85	x	x	x	x	x	x	x	x	x
Biogas (400kW)[1]	12.50	x	x	x	x	x	x	x	x	x
Electricity system (price based on 1 kW-100 sys)[2]	15.68	x	x	x	x	x	x	x	x	x
Electricity system (price based on 1 kW-1000 sys)[2]	9.00	x	x	x	x	x	x	x	x	x
Bloom 100kW system[3]	16.92	x	x	x	x	x	x	x	x	x
Elec system - 270 kW - 50 units installed[4]	2.15	x	x	x	x	x	x	x	x	x
PEMFC - 80 kW[5]	4.91	x	x							
Auxillary power - 1 kW - 1 unit[6]	9.84	x	x	x	x	x	x	x	x	x
Auxillary power - 1 kW - 100 units[6]	8.64	x	x	x	x	x	x	x	x	x
Auxillary power - 5 kW - 1 unit[6]	3.10	x	x	x	x	x	x	x	x	x
Auxillary power - 5 kW - 100 units[6]	2.52	x	x	x	x	x	x	x	x	x
Stationary hybrid SOFC - 3.1 MW (2.6 MW fuel cell)[7]	2.02	x	x	x	x	x				

Sources: [1](Ammermann et al., 2015), include installation, are based on a 'generic' system so not one specific type of fuel cell, but one that fits the purpose [2](James et al., 2012) [3](Mcphail et al., 2013) [4](Weimar et al., 2013) [5](Schoots et al., 2010) [6](Battelle, 2014) [7](Thijssen, 2007)

Appendix H – Learning rates and installed capacity

Learning rate Fuel cell

Learning rates about fuel cells and SOFC more specifically were gathered and are shown in Table 51. Rivera-Tinoco, Schoots, & van der Zwaan (2012) were the first to do an extensive learning rate analysis specifically for SOFC-cells. They focused on planar SOFCs, and not on the whole system but on the cells itself. They divided the learning rate both into learning in different stages of SOFC development and distinguish different types of learning. For the learning curve analysis in this research, the price is based on an early commercial system (Ammerman et al., 2015), and all possible types of learning can be taken into account. The learning rate given by Rivera-Tinoco et al. (2012) for early commercial systems is 12%. As most technologies the learning rate have a learning rate between 5%-30% (Blok, 2007), this is a reasonable number.

Other sources show somewhat higher numbers see Table 51. The number of Schoots, Kramer, & van der Zwaan (2010) is especially for PEMFC-cells (Proton Exchange Membrane Fuel Cell) and not for SOFC. Weimar, Gotthold, Chick, & Whyatt (2013) have adopted their learning rate to aerospace processes as they are focused on 5kW systems for auxiliary power systems, which is not that applicable for the current research. The NETL report (Iyengar et al., 2013) focuses on SOFC stacks, but not a complete system, and states that their numbers are (partly) based on (Rivera-Tinoco et al., 2012). Finally Thijssen comes up with a 10-20% learning rate for SOFC systems in the period 2020-2030 without a very clear foundation for this number.

So for only the SOFC stack itself, a learning rate between 12 %– 20% seems plausible. However, the article or Rivera-Tinoco et al. (2012) appears to be the most reliable source and reports a 12% learning rate. On the other hand, it does not cover the whole SOFC system but only the SOFC stack itself. Though for instance the stack packaging is likely to be reduced considerably by economics of scale effects (Thijssen, 2007), which is not included in the 12% learning rate. To include these effects to some extent, in the end a 15% learning rate was chosen.

Table 51 Learning of SOFCs. Learning rates from different stages, applications and including different factors. eco-of-scale = economies of scale.

Fuel cell learning	Learning rate (lr)	Stage/ application	pure learning	eco-of-scale material	eco-of-scale equipment	eco-of-scale auto-mation	Learning + auto-mation	Application
(Rivera-Tinoco et al., 2012)	16%	R&D	16%	16%	16%	16%	SOFC - general
	44%	Pilot	27%	44%	28%	36%	SOFC - general
	12%	**Early commercial**	1%	5%	12%	10%	SOFC - general
	35%	All stages	20%	27%	22%	28%	SOFC - general
(Schoots et al., 2010)	21%		x					PEMFC
(Weimar et al., 2013)	15%	R&D	x					SOFC – 5kW for auxillary applications
(Iyengar et al., 2013)	20%		x	x	x	x		SOFC, distributed generation
(Thijssen, 2009)	10-20%		?					SOFC, general

Global installed capacity SOFC

For the cumulative global installed capacity it was hard to find enough recent data, but more data were found on global SOFC shipments. We assumed that shipments correlate well enough with global installed capacity to use it as a proxy.

Data from global shipments from 2009-2015 were taken from (Carter & Wing, 2013) and (Hart et al., 2015). These data added up to 194 MW, global shipments before 2009 were assumed to be low enough to be neglected.

Table 52 Global SOFC shipments, data taken from (Carter & Wing, 2013; Hart et al., 2015).

SOFC Global shipments in MW	
2009	1.1
2010	6.7
2011	10.6
2012	26.9
2013	47
2014	38.2
2015	63.1
Sum	194

Gasifier system

Aranda et al. (2014) reported a learning rate of 10% for a complete biomass to SNG system and applied it on a 1GW plant with expected installed capacity of 10MW in 2030. A 14% learning rate was adopted by van den Broek et al. (2009) for the capital investment of the gasifier area of an Integrated Gasification Combined Cycle (IGCC). The installed capacity value here was 10 GW, but based on 2001 data. Finally (Knoope et al., 2013) reported an average learning rate of 6% for the gasifier. The total installed IGCC capacity based on biomass is 0 GW and expected to be 0.4 GW in 2030, coal based IGCC is 4.3 GW and expected to grow to 13 GGW in 2020. Installed capacity (in 2010) of just a gasifier and cleaning system was 99 GWth, and expected to increase to 189 GWth in 2020, though these values are not specifically for biomass gasification.

In this study a biomass gasification system is considered that does not seem to exist at very large scale. Although the data are not very clear, a 10 GW value was adopted for the current calculations. That is not too small, but also not too optimistic.

For the learning rate, the average value of 10% was used.

Appendix I – Total results of sensitivity analysis

Table 53 Sensitivity range and results for amount of toilet paper on LCOE

	Amount of toilet paper	25,000
		WTP on LCOE
		0.2305
80%	20000	0.2656
82%	20600	0.2605
85%	21200	0.2557
87%	21800	0.2512
90%	22400	0.2469
92%	23000	0.2428
94%	23600	0.2389
97%	24200	0.2352
99%	24800	0.2317
102%	25400	0.2283
104%	26000	0.2250
106%	26600	0.2220
109%	27200	0.2190
111%	27800	0.2161
114%	28400	0.2134
116%	29000	0.2108
118%	29600	0.2083
120%	30000	0.2067

Table 54 Sensitivity range and results for water content on LCOE

Water content now	60%	
		Water content on LCOE
		0.2305
83%	50%	0.2018
87%	52%	0.2067
90%	54%	0.2119
93%	56%	0.2176
97%	58%	0.2237
100%	60%	0.2305
103%	62%	0.2380
107%	64%	0.2462
110%	66%	0.2553
113%	68%	0.2656
117%	70%	0.2772

Table 55 Sensitivity range and results for SOFC efficiency on LCOE

	Efficiency	55%
		Efficiency fuel cell on LCOE
		0.2305
81.82%	45%	0.2817
90.91%	50%	0.2536
100.00%	55%	0.2305
109.09%	60%	0.2113
118.18%	65%	0.1950
127.27%	70%	0.1811

Table 56 Sensitivity range and results for WTP price on LCOE

WTP price now (euro/tonne)	0
	WTP price on LCOE
	0.2305
-70	**0.1359**
-60	0.1494
-50	0.1630
-40	0.1765
-30	0.1900
-20	0.2035
0	0.2305
20	**0.2575**
40	0.2846

Table 57 Sensitivity range and results for the SOFC investment cost on LCOE

	SOFC cost	€ 11,850,583
		SOFC price on LCOE
		0.2305
21%	€ 2,500,000	0.1770
34%	€ 4,000,000	0.1855
42%	€ 5,000,000	0.1913
51%	€ 6,000,000	0.1970
59%	€ 7,000,000	0.2027
67%	€ 8,000,000	0.2084
76%	€ 9,000,000	0.2142
84%	€ 10,000,000	0.2199
93%	€ 11,000,000	0.2256
110%	€ 13,000,000	0.2371
127%	€ 15,000,000	0.2485
143%	€ 17,000,000	0.2600

Table 58 Sensitivity range and results for the gasifier + cleaning stystem OM cost on LCOE

	OM cost - gasifier	€ 713,064.00
		OM cost on LCOE
		0.2305
77%	€ 550,000	0.2224
84%	€ 600,000	0.2249
91%	€ 650,000	0.2274
98%	€ 700,000	0.2299
105%	€ 750,000	0.2323
112%	€ 800,000	0.2348
115%	€ 820,000	0.2358
115%	€ 820,000	0.2358
115%	€ 820,000	0.2358
115%	€ 820,000	0.2358
115%	€ 820,000	0.2358
119%	€ 850,000	0.2373
126%	€ 900,000	0.2398
133%	€ 950,000	0.2422
154%	€ 1,100,000	0.2497

Table 59 Sensitivity range and results for the SOFC OM cost on LCOE

	OM cost - SOFC	€ 198,393.91
		OM cost on LCOE
		0.2305
81%	€ 160,000	0.2286
86%	€ 170,000	0.2291
91%	€ 180,000	0.2296
96%	€ 190,000	0.2301
101%	€ 200,000	0.2306
106%	€ 210,000	0.2311
111%	€ 220,000	0.2316
116%	€ 230,000	0.2321
121%	€ 240,000	0.2326
126%	€ 250,000	0.2331
131%	€ 260,000	0.2336
136%	€ 270,000	0.2341
141%	€ 280,000	0.2345
146%	€ 290,000	0.2350
151%	€ 300,000	0.2355

Table 60 Sensitivity range and results for the discount rate on LCOE

	Discount rate	5%
		Discount rate on LCOE
		0.2305
60.0%	3%	0.2100
80.0%	4%	0.2201
100.0%	5%	0.2305
120.0%	6%	0.2413
140.0%	7%	0.2524
160.0%	8%	0.2638
180.0%	9%	0.2756
190.0%	10%	0.2815
196.0%	10%	0.2851
200.0%	10%	0.2876

Table 61 Sensitivity range and results for the SOFC scale factor on LCOE

	Scale factor	0.85
		Scale factor on LCOE
		0.2305
94%	0.8	0.2270
96%	0.82	0.2284
99%	0.84	0.2298
101%	0.86	0.2312
104%	0.88	0.2327
106%	0.9	0.2342
108%	0.92	0.2357
112%	0.95	0.2379

Table 62 Sensitivity range and results for the project life time on LCOE

Lifetime on LCOE			
	Project lifetime	LCOE	
63%	10	0.2916	
94%	15	0.2387	
100%	16	0.2305	
125%	20	0.2225	
156%	25	0.2067	
188%	30	0.19	

Appendix J – Pedigree matrices and diagnostic diagram table

Table 63 Pedigree matrix filled in by Chris Reijken from Waternet

	Proxy	Reliability of source	Completeness	Validation process
Amount of toilet paper	2	2	0	1
Water content toilet paper	2	1	0	2
WTP composition	2	1	1	0
Price of WTP	2	3	2	3

Table 64 Pedigree matrix filled in by supervisor, pHd candidate at the Energy and Resources group (Copernicus Institute) at Utrecht University.

	Proxy	Reliability of source	Completeness	Validation process
Amount of toilet paper	4	1	0	1
Water content toilet paper	4	1	2	1
WTP composition	3	4	3	2
Price of WTP	4	1	2	0
SOFC investment costs	3	1	2	3
Efficiency SOFC	3	2	3	2
Scale factor SOFC	2	2	1	0
Learning rate SOFC	3	2	1	1
Project Lifetime	4	2	3	2
OM costs vergasser	4	4	3	2
OM costs FC	4	1	2	0
Subsidy value	3	4	4	2
Discount rate	4	4	2	3

Table 65 Pedigree matrix filled in by Robin Zwart, technology and product development manager at Royal Dahlman.

	Proxy	Reliability of source	Completeness	Validation process
Amount of toilet paper	3	4	0	-
Project Lifetime	2	1	0	2
OM costs vergasser	2	3	3	2
Subsidy value	2	4	3	2
Discount rate	2	-	-	-

Table 66 Pedigree matrix filled in by Wouter Schakel, researcher at the Energy and Resources group (Copernicus Institure) at Utrecht University.

	Proxy	Reliability of source	Completeness	Validation process
Amount of toilet paper	4	1	0	1
Water content toilet paper	4	1	1	1
WTP composition	3	4	2	1
Price of WTP	4	0	0	0
SOFC investment costs	2	1	2	0
Efficiency SOFC	3	1	2	2
Scale factor SOFC	3	1	0	0

	Proxy	Reliability of source	Completeness	Validation process
Learning rate SOFC	3	2	1	2
Project Lifetime	4	1	2	1
OM costs (Gasifier)	3	3	3	1
OM costs (SOFC)	3	1	1	1
Subsidy Value	1	4	2	2
Discount rate	4	1	1	3

Table 67 Pedigree matrix filled in by Cora Fernandez Dacosta, researcher at the Energy and Resources group (Copernicus Institute) at Utrecht University.

	Proxy	Reliability of source	Completeness	Validation process
Amount of toilet paper	4	3	2	0
Water content toilet paper	3	3	3	2
WTP composition	4	4	4	0
Price of WTP	4	4	3	0
SOFC investment costs	2	2	2	1
Efficiency SOFC	3	2	2	2
Scale factor SOFC	3	1	2	0
Learning rate SOFC	3	2	3	3
Project Lifetime	3	3	3	3
OM costs (Gasifier)	4	4	4	2
OM costs (SOFC)	2	2	2	1
Subsidy Value	2	1	2	1
Discount rate	3	3	3	3

Table 68 Pedigree matrix filled in by the author

	Proxy	Reliability of source	Completeness	Validation process
Water content toilet paper	4	2	1	1
Amount of toilet paper	3	0	1	1
WTP composition	4	3	3	2
Price of WTP	4	3	2	2
SOFC investment costs	3	2	4	3
Efficiency SOFC	4	1	3	2
Scale factor SOFC	4	0	0	0
Learning rate SOFC	2	2	3	3
Project Lifetime	4	3	2	3
OM costs - gasifier	3	4	3	2
OM cost - SOFC	3	1	1	0
Subsidy value	3	1	2	0
Discount rate	4	4	4	3

Table 69 Table used as input for the diagnostic diagram. The average of the median scores is given as values for the x-axes, while the normalized sensitivity values were used for the y-axes.

	Average - Med	Sensitivity (average deviation from LCOE in euro/kWh)	Normalized sensitivity
Amount of toilet paper	1.50	2.95	0.39
Water content toilet paper	1.75	3.77	0.50
WTP composition	2.75	0.26	0.04
Price of WTP	2.25	7.43	0.99
SOFC investment costs	2.00	4.15	0.55
Efficiency SOFC	2.25	5.03	0.67
Scale factor SOFC	1.13	0.55	0.07
Learning rate SOFC	2.38	1.53	0.20
Project Lifetime	2.50	4.91	0.66
OM costs- Gasifier	3.00	1.36	0.18
OM cost - SOFC	1.50	0.35	0.05
Subsidy value	2.50	7.50	1.00
Discount rate	3.25	3.88	0.52