Energy research Centre of the Netherlands

High efficiency power production from biomass and waste

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Acknowledgement/Preface

This report presents results of the project "High efficiency power production from biomass and waste, proof of concept" executed by the ECN unit "Biomass, Coal & Environmental Research" in co-operation with TechnoInvent. Part of the work at ECN was carried out by the units "Hydrogen & Clean Fossil Fuels" and "Engineering & Services". The project received financial support from the BSE programme managed by SenterNovem under contract number 2020-03-11-14-010.

Abstract

Two-stage gasification allows power production from biomass and waste with high efficiency. The process involves pyrolysis at about 550°C followed by heating of the pyrolysis gas to about 1300°C in order to crack hydrocarbons and obtain syngas, a mixture of H_2, CO, H_2O and CO_2. The second stage produces soot as unwanted by-product. Experimental results are reported on the suppression of soot formation in the second stage for two different fuels: beech wood pellets and Rofire pellets, made from rejects of paper recycling. Syngas obtained from these two fuels and from an industrial waste fuel has been cleaned and fed to a commercial SOFC stack for 250 hours in total. The SOFC stack showed comparable performance on real and synthetic syngas and no signs of accelerated degradation in performance over these tests.

The experimental results have been used for the design and analysis of a future 25 MW_{th} demonstration plant. As an alternative, a 2.6 MW_{th} system was considered which uses the Green MoDem approach to convert waste fuel into bio-oil and syngas. The 25 MW_{th} system can reach high efficiency only if char produced in the pyrolysis step is converted into additional syngas by steam gasification, and if SOFC off-gas and system waste heat are used in a steam bottoming cycle for additional power production. A net electrical efficiency of 38% is predicted. In addition, heat can be delivered with 37% efficiency. The 2.6 MW_{th} system with only a dual fuel engine to burn bio-oil and syngas promises nearly 40% electrical efficiency plus 41% efficiency for heat production. If syngas is fed to an SOFC system and off-gas and bio-oil to a dual fuel engine, the electrical efficiency can rise to 45%. However, the efficiency for heat production drops to 15%, as waste heat from the SOFC system cannot be used effectively.

The economic analysis makes clear that at -20 €/tonne fuel, 70 €/MWh for electricity and 7 €/GJ for heat the 25 MW_{th} system is not economically viable at the price level of 1600 €/kW_e expected for SOFC systems in the near future and 200 €/kW_{th} for two stage gasification and gas cleaning. In order to reach break-even, the electricity price would have to rise to 141 €/MWh. For break-even at 70 €/MWh, the price of both main parts would have to drop to e.g. 500 €/kW_e for the SOFC and to 80 €/kW_{th} for the gasification and gas cleaning. At the higher equipment price level, break-even may already be reached at an electricity price of 87 €/MWh if more favourable conditions apply, i.e. -100 €/tonne fuel, CO_2 credits of 10 €/tonne, cheap labour and full use of the available heat. In that case, the simple pay-back time would become 9.3 years. In those favourable conditions, the simple pay-back time of the base case Green MoDem system with dual-fuel engine would be 4.9 years. Addition of an SOFC system would increase the simple pay-back time to 10.5 years.

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Summary

A previous study showed that two-stage gasification allows power production from biomass and waste with high efficiency. The process involves pyrolysis at about 550°C followed by heating of the pyrolysis gas to about 1300°C in order to crack hydrocarbons and obtain syngas, a mixture of H_2, CO, H_2O and CO_2. Cleaned syngas can be converted into electricity by an SOFC system. Off-gas from the SOFC and system waste heat can supply heat for steam generation and additional power production by a steam turbine. According to the previous study, the net electrical efficiency could be increased even further if syngas would be cleaned at high temperature. The aims of the present study were to prove the concept of two-stage gasification plus SOFC at 5 kW_e scale, design and test hot syngas cleaning, and design a demonstration plant that could serve as the next step in the development.

Soon after the start of the project it became clear that a 5 kW_e SOFC system could not be realized. The partner responsible for the SOFC system withdrew from the project. A different supplier was found, who offered 500 W_e SOFC stacks for assembly into an SOFC system by ECN. Eventually, the stack performance proved to be only 350 W_e. However, the step from a laboratory stack of 5 cells to a commercial stack of 30 cells still fits the principle of scaling from the feasibility stage to the concept stage. Other scaling factors are the duration of the tests, which was increased from 50 hours to 250 hours, and operation of an integral system instead of a system with intermediary gas storage.

The main problem encountered in the previous study was the production of soot which blocks gas tubes. Much effort was spent on the reduction of soot formation. Steam addition proved effective, but significantly more steam is needed for waste fuel containing plastics than for clean wood fuel. In order to remove residual soot and allow continuous operation, the test facility was equipped with a set of two hot gas filters which can be used and cleaned alternately. Reactors with commercial adsorbents were added for the removal of S and Cl compounds.

The commercial SOFC stack was mounted in an ECN test stand. The integral system of two-stage gasification, hot gas cleaning and SOFC system was used for tests with three fuels for a total of 250 hours. The fuels used were clean beech wood, Rofire pellets made from paper recycling rejects and industrial waste. Although some problems were encountered, especially during the first test with clean wood and after the test with Rofire pellets, the tests were very successful. The stack performance on syngas was comparable with the performance on synthetic syngas and showed no signs of accelerated degradation.

The experimental results have been used for the design and analysis of a future 25 MW_{th} demonstration plant. The system can reach high efficiency only if char produced in the pyrolysis step (i.e. the first gasification stage) is converted into additional syngas by steam gasification, and if SOFC off-gas and system waste heat are used in a steam bottoming cycle for additional power production. A net electrical efficiency of 38% is predicted. In addition, heat can be delivered with 37% efficiency.

Project partner TechnoInvent has analysed a 2.6 MW_{th} system based on its Green MoDem technology to convert waste fuel into bio-oil and syngas. Such a system with only a dual fuel engine to burn bio-oil and syngas promises nearly 40% electrical efficiency plus 41% efficiency for heat production. If syngas is fed to an SOFC system and off-gas and bio-oil to a dual fuel engine, the electrical efficiency can rise to 45%. However, the efficiency for heat production drops to 15%, as waste heat from the SOFC system cannot be used effectively.

The economic analysis makes clear that at -20 €/tonne fuel, 70 €/MWh for electricity and 7 €/GJ for heat the 25 MW_{th} system is not economically viable at the price level of 1600 €/kW_e expected for SOFC systems in the near future and 200 €/kW_{th} for two stage gasification and gas cleaning. In order to reach break-even, the electricity price would have to rise to 141 €/MWh. For break-even at 70 €/MWh, the price of both main parts would have to drop to e.g. 500 €/kW_e for the SOFC and to 80 €/kW_{th} for the gasification and gas cleaning. At the higher equipment price level, break-even may already be reached at an electricity price of 87 €/MWh if more favourable conditions apply, i.e. -100 €/tonne fuel, CO_2 credits of 10 €/tonne, cheap labour and full use of the available heat. In that case, the simple pay-back time would become 9.3 years.

In those favourable conditions, the simple pay-back time of the base case Green MoDem system with dual-fuel engine would be 4.9 years. Addition of an SOFC system would increase the simple pay-back time to 10.5 years.

For both systems, the high electrical efficiency of SOFC systems does not suffice to pay the price expected in the near future for SOFC systems. On the other hand, even the largest SOFC system in development will be a factor 4 smaller than needed for the Green MoDem system. Some development at that front will be needed too. If in the mean time the electricity price will keep rising, and if the goals of cost reduction for SOFC systems will be met, there may still be a future for two-stage gasification combined with SOFC systems.

1. Introduction

The production of energy from biomass and waste contributes to the reduction of CO_2 emissions. In the Netherlands, biomass and waste combustion is applied at large scale in municipal waste incinerator (AVIs) and coal power plants. The electrical efficiency lies typically between 20% and 25% for AVIs and around 40% for coal power plants. With use of available new technology and more expensive materials these figures can rise to 30% and 45% respectively. In future, high temperature solid oxide fuel cells (SOFCs) promise 40% to 50% efficiency, to be reached in relatively small scale installations. The relatively small scale makes it easier to utilise waste heat and thus further increase the total system efficiency.

For use in fuel cells, biomass and waste have to be converted into clean combustible gas. We investigate the suitability of a two-stage gasification process and hot gas cleaning for that purpose. The first stage of the gasification process is slow pyrolysis. Dry biomass or waste is heated to about 550°C in inert atmosphere. In those conditions, organic material is converted to pyrolysis gas and char. Inorganic materials end up predominantly in the char.

Char can be burned to provide heat for the gasification process or other purposes, e.g. steam production for additional power generation. Char can also be converted by steam gasification into producer gas. If waste is used, inorganic contaminants will be concentrated in the char and in flue gas from char combustion. If suitable waste streams and additives are combined, char can be used as an energy source to melt the inorganic contaminants and immobilize them in a valuable artificial basalt product.

Pyrolysis gas contains mainly CO, CO_2, H_2, H_2O, CH_4 and other compounds of C, H and O. A large fraction of the pyrolysis gas energy content stems from heavy hydrocarbons. These cannot be applied in SOFCs because of their tendency to soot formation. In the second stage of the gasification process, pyrolysis gas is heated in a thermal tar cracker to about 1300°C to obtain syngas, a mixture of CO, CO_2, H_2 and H_2O. After removal of contaminants, the cleaned syngas can be applied in SOFCs.

The conversion of hydrocarbons to syngas at high temperatures yields soot as a by-product. One of the project aims is to find practical operating conditions that minimise soot formation to the extent that dust filters can be used effectively to remove any remaining soot.

When biomass is used as fuel, gaseous contaminants will be limited to small amounts of H_2S and HCl. In case of waste streams, more significant amounts and other contaminants may be present, e.g. Cl from PVC and volatile metals or metal compounds from electronic scrap. In the present project we consider only the removal of H_2S and HCl.

In a previous project, syngas cleaning was performed at room temperature [1]. That meant, syngas had to be cooled from above 1000°C to room temperature and reheated to 750°C before use in the fuel cell. One of the aims of the present project is to show the feasibility of syngas cleaning at elevated temperature. That approach limits the cooling and heating loads and increases the system efficiency.

In the previous project, we demonstrated the operation of an experimental 5-cell SOFC stack with clean syngas obtained by two-stage gasification. The present project was intended as a proof of concept with a commercial SOFC system. The performance was to be monitored over three periods of 100 hours, using clean biomass and two biomass-based waste streams. Results were to be used for the design of a demonstration project and analysis of the expected performance and economics of a commercial system.

In the original project plan the use of a commercial 5 kW_e SOFC system was announced. However, financial problems arose which forced us to find a different supplier and limit our ambitions. Eventually, a commercial 30-cell 300 W_e SOFC stack was used. The discussions about the financial problems and the search for an alternative supplier caused nearly two years delay in the execution of the project.

The original project plan referred to a waste treatment facility to be built by a project partner as future site for a demonstration of the SOFC technology. However, the facility has not been realized and the partner involved has transferred its project tasks to another party with approval of SenterNovem. This change of project partners has also led to the introduction of the "green modem" technology, as an alternative for two-stage gasification, and a slightly different view on the use of char from the slow pyrolysis process.

The following chapters describe the work performed within the project. The second chapter discusses the design and layout of the test installation and experimental procedures. The third chapter describes the fuels used and results of optimization experiments. The fourth chapter gives results of the endurance tests of the total system including the SOFC stack. The fifth chapter contains a model analysis of the system used and an analysis of various options to enhance the efficiency. More details are given in appendix A. The sixth chapter contains a description of two designs, for 2.5 and 25 MW scale respectively, and an analysis of their expected economical performance. The final chapter summarizes results and conclusions.

2. Experimental installation and procedures

This chapter describes the experimental installations at ECN used for the optimization and endurance experiments. The chapter is divided into sections which treat a specific part of the installation with operating procedures and standard conditions. The first section describes the existing Pyromaat installation for two-stage gasification and modifications made to it in the course of the project. The second section discusses the design and layout of the gas cleaning. The third section describes the SOFC test set-up and the fourth section the gas analysis.

2.1 Two-stage gasification installation Pyromaat

2.1.1 Existing system

A schematic drawing of the two-stage gasification installation Pyromaat as used in the previous project is shown in Figure 2.1. A feeding screw in the feed bunker delivers biomass or waste fuel to the externally heated pyrolysis reactor. There, the fuel is heated to about 500°C and converted into pyrolysis gas and char. Solid material is moved through the pyrolysis reactor by a slowly rotating screw. The rotation speed is usually set at a value which propels material in 15 minutes to the char collection point.

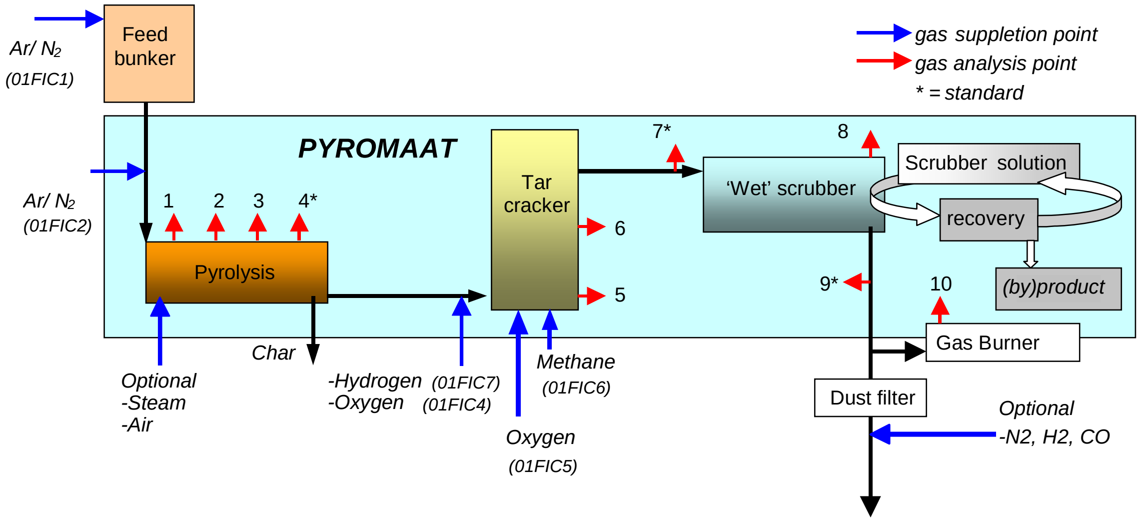

Figure 2.1 *Scheme of existing Pyromaat two-stage gasification set-up.*

In the tar cracker, the pyrolysis gas is broken down at about 1300°C to a fuel gas mixture of CO, CO_2, H_2 and H_2O. The tar cracker is heated internally by a methane oxygen flame. At a number of points Ar or N_2 can be added as carrier, tracer or buffer gas or gas samples can be taken for analysis. Downstream of the tar cracker the fuel gas is cooled and partially cleaned by a wet scrubber. The fuel gas is then burnt in a gas burner or cleaned more thoroughly for advanced applications.

2.1.2 Modifications

The present project is to realise a direct coupling between the Pyromaat and an SOFC system and increase the system efficiency by gas cleaning at high temperature. To that end, fuel gas has to be taken directly from the tar cracker and cooled only to the maximum operating temperature of the gas cleaning. Actually, given the modest gas flow required, in our installation there is no need for active gas cooling. On the contrary, heat loss had to be limited by trace heating of gas lines. The new gas cleaning and SOFC systems are described in the next sections. Figure 2.2 shows the modified system and the added parts. Double lines in the figure indicate that gas lines and reactors are insulated and heated.

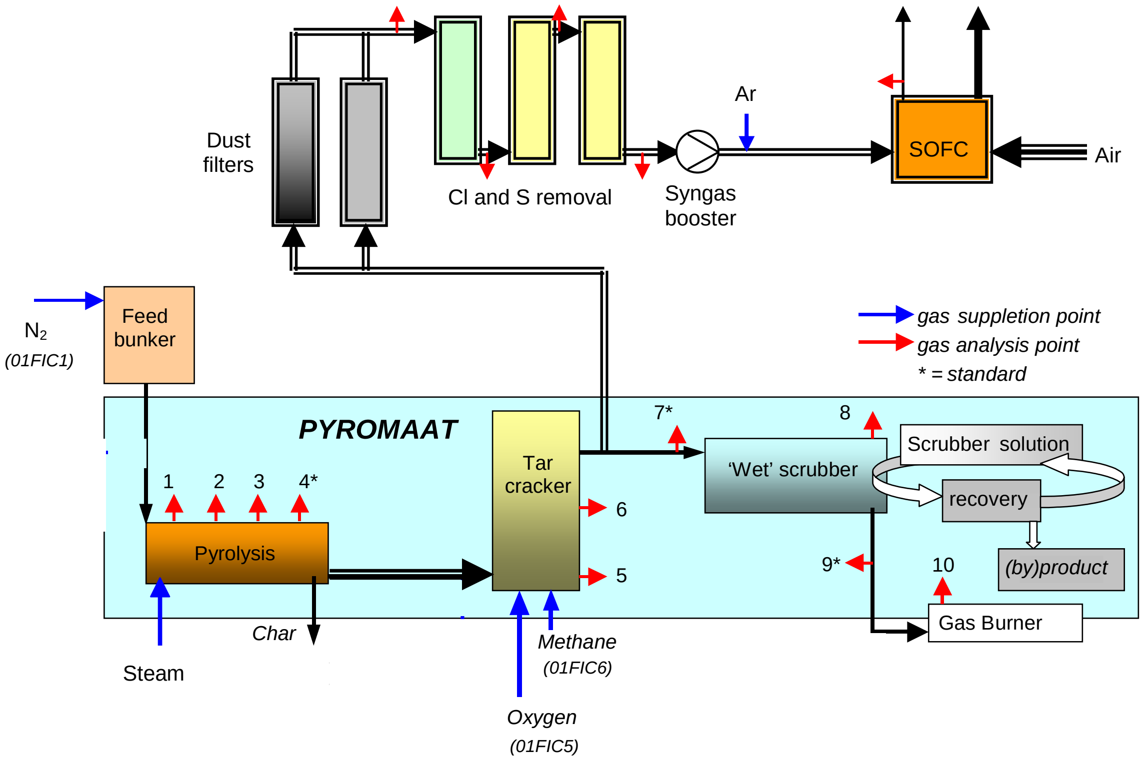

Figure 2.2 *Scheme of modified Pyromaat two-stage gasification set-up, gas cleaning and SOFC system.*

In the existing set-up, the wet scrubber removed most of the soot produced in the tar cracker. In order to limit the dust load on the new hot gas cleaning, much effort was spent on suppression of soot formation. To that end, the steam injection capacity at the pyrolysis reactor was increased from 0.5 to 3 kg/hr. The increased amount of steam replaces the Ar used previously as carrier gas in the pyrolysis reactor.

In order to be able to monitor the fuel feed, the feed bunker was equipped with an on-line weighing unit. Unfortunately, the weighing signal showed oscillations with an amplitude similar in size to the hourly feed rate. The oscillations are caused by the feed screws in the fuel feed line and in the pyrolysis reactor. Due to the coupling of the bunker with the feed line and the pyrolysis reactor, any asymmetry of the feed screws translates into vertical movement of the feed bunker. Attempts to monitor the fuel feed on-line were abandoned. Average feed rates given in this report are derived from the net amount of fuel fed over the duration of a test.

In a commercial system, the pyrolysis reactor would be designed to deliver gas at sufficient pressure to overcome pressure losses in the gas cleaning, gas lines and SOFC system. As the Pyromaat at ECN is insufficiently gas-tight, we have to operate it at atmospheric pressure and use a booster pump downstream of the gas cleaning to deliver gas to the SOFC system. The use of a booster pump also enables us to take just the amount of gas needed. A small Ar flow is added downstream of the booster pump to allow monitoring of the fuel gas flow to the SOFC system. Surplus fuel gas is still burnt in the gas burner shown.

2.1.3 Operating procedures

The fuel feed rate depends on the rotation speed of the fuel feeding screw and the fuel properties. It may also vary with the filling degree of the fuel bunker. The relation between screw rotation speed and feed rate is determined beforehand for each fuel. The average fuel feed rate is also calculated afterwards from the amount of fuel used and the operating time.

The feed bunker has a capacity sufficient for a few hours of operation. For longer runs, the bunker can be refilled through a system of valves. Usually, a fixed amount is added, corresponding to about one hour and a half of operation. The amount of fuel added is noted in a log file. At the end of a test run, the fuel bunker is emptied and its contents weighed to allow calculation of the average fuel feed rate.

A small constant flow of N_2 is added to the fuel bunker to keep it free of pyrolysis gas. In the course of the tests it proved necessary to increase the N_2 flow to keep steam away from the fuel. Otherwise, fuel stuck to the walls and damaged the feeding screw.

Char is collected in metal buckets which are replaced about every two hours. The used bucket is provisionally sealed from air and put into a water bath for cooling. After cooling, the bucket is dried and weighed to determine its char content. The char weight is noted in a log file.

Both when fuel is added to the feed bunker and when char collection buckets are exchanged, care is taken to prevent air from leaking into the pyrolysis reactor or pyrolysis gas from leaking to the room. To that end, gas volumes are flushed with N_2 before valves are opened. For personal protection, operators wear individual CO monitors. In addition, fixed CO detectors monitor the working environment.

In preparation of and in between experiments, the tar cracker is heated by combustion of natural gas with O_2. The gas flows are switched off if no flame UV signal is detected. When the tar cracker is cold, it takes about two days to warm it throughout. During experiments, natural gas is replaced by CH_4. The CH_4 and O_2 flows are controlled by calibrated mass flow controllers. For safety reasons, the system does not allow to switch on O_2 unless there is a flow of CH_4.

2.2 Gas cleaning equipment

The SOFC system uses only part of the total gas flow from the tar cracker. This section describes the new gas cleaning installed to allow SOFC operation. As mentioned before, surplus gas is burnt in the gas burner shown in Figure 2.2. The existing wet scrubber has been used only to treat surplus gas when the fuel contained significant amounts of Cl.

Figure 2.3 *Set of hot gas filters with temperature control unit (left). The orange handles allow opening and closing of high-temperature valves for switching between filters.*

2.2.1 Dust removal

Fuel gas from the tar cracker is cleaned in three steps. First, soot and dust are removed by ceramic filter candles. The material can be used up to 600°C. As temperature has no effect on the filter efficiency, we chose to keep the actual operating temperature to 300°C to reduce the trace heating duty. Two filter holders have been installed (see Figure 2.3) with provisions to allow switching from one unit to the other without interruption of the endurance test. Each filter holder can be isolated from the system to allow demounting and cleaning or replacement of its filter candle.

2.2.2 Cl and S removal

The second gas cleaning step is chlorine removal. We use a commercial absorbent with Na_2O as main active component. According to specifications, the maximum operating temperature of the material we used is limited to about 250°C. We kept the temperature at 200°C. However, similar materials are available for use at 400°C and more. Our supplier confirmed that the temperature limitation is not caused by the active ingredient but by other components.

The third gas cleaning step is sulfur removal. We use a commercial absorbent with ZnO as main active component. The material can be used up to 400°C, but the residual sulfur concentration is expected to increase with operating temperature. We chose to use the same temperature as for the Cl removal. In order not to jeopardize the endurance tests, we use a sulfur guard bed downstream of the high-temperature gas cleaning for additional protection of the SOFC stack. Figure 2.4 shows a picture of the three vessels with absorbents.

Figure 2.4 *Set of three vessels with absorbents for removal of Cl and S from gas.*

2.3 SOFC test set-up

2.3.1 Test facility

SOFC stacks containing 30 cells with metallic interconnects were acquired from Staxera. Each cell has an active area of 81 cm^2. The nominal power output of a stack is 300 W for operation on H_2 and air. The SOFC system was mounted in a purpose-built oven to limit heat loss and give control over the stack temperature. Staxera supplied some of the components needed and technical drawings of other parts. Figure 2.5 and Figure 2.6 show the fuel cell stack in the oven. The oven with fuel cell stack is mounted in a fuel cell test rig (see Figure 2.7).

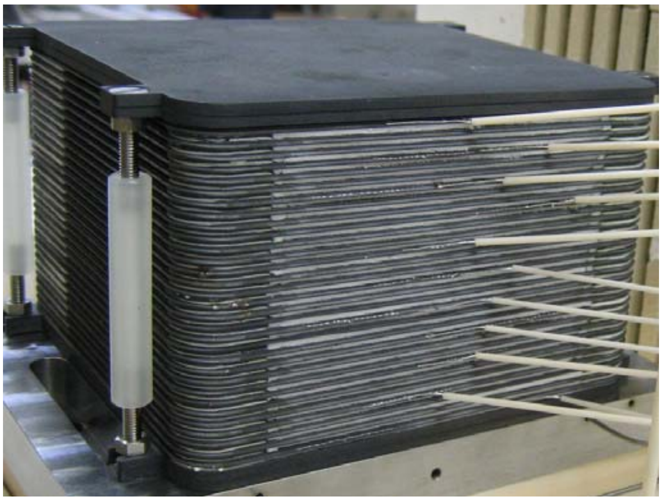

Figure 2.5 *Staxera SOFC 30 cell stack with connections for voltage measurements at every 3th cell.*

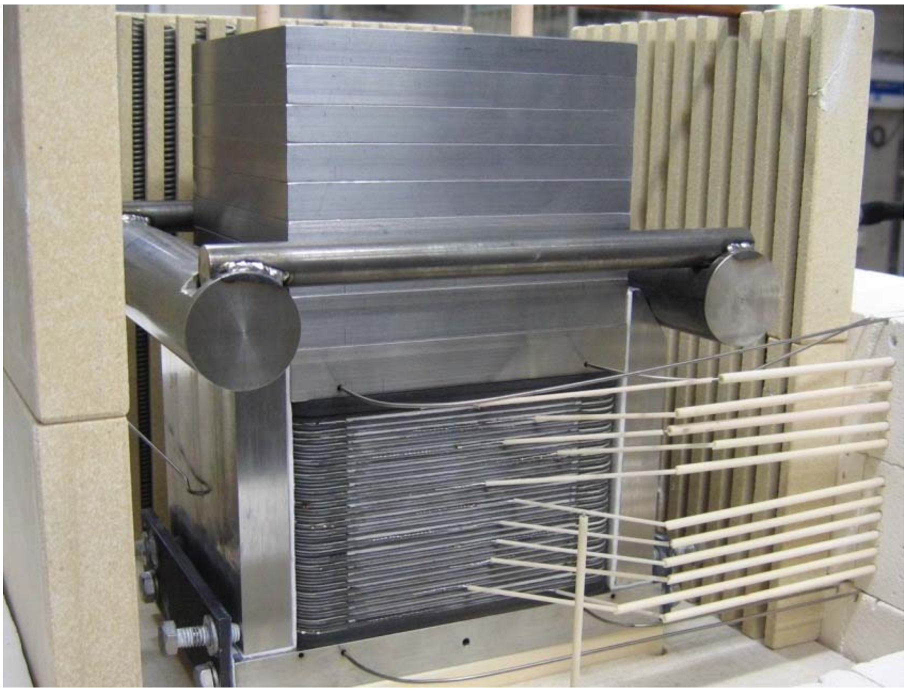

Figure 2.6 *Staxera SOFC 30 cell stack with air supply hoods (left and right) and weights (top) to keep pressure on gas seals and active area.*

Figure 2.7 *SOFC test facility with load unit and computer (left), furnace with gas pre-heaters (centre), fuel gas supply (right) and off-gas ducts (top).*

2.3.2 Test conditions

The SOFC stack is operated at a temperature of 800 to 850°C. The stack is mounted in an oven which guarantees the temperature of the surrounding. Fuel gas and air are preheated to about 750°C and leave the stack at about 850°C. Thermocouples were mounted in the air and gas supply and off-gas lines. Unfortunately, the thermocouples in the air hoods ceased functioning in the first test run, so the stack temperature had to be derived from measurements of four thermocouples mounted in the top and base plates (see Figure 2.6).

The SOFC stack is flushed with 0.6 Nm^3/hr of a mixture containing 40% H_2 and 60% N_2 when not in operation with syngas. During syngas tests, the gas flow is set to a value which yields 60% to 65% fuel utilization. The stack current is kept a 10 A by a load unit. The condition of the stack is monitored regularly by IV and impedance measurements. During those measurements, the stack current is varied.

The stack is supplied with contacts for voltage measurements over ten sets of three cells (see Figure 2.5). All temperatures, voltage measurements and the load current are registered and stored automatically in a database.

2.4 Gas analysis

Gas samples for analysis can be taken at the pyrolysis unit, downstream of the tar cracker, after each step in the gas cleaning, downstream of the booster pump and Ar addition point and downstream of the SOFC system. The equipment and procedures used at various points are described below.

Pyrolysis gas

The composition of dry pyrolysis gas was determined only for a few short periods during the endurance tests. The composition of "untreated" pyrolysis gas is difficult to measure, as tar compounds quickly pollute the gas lines. As the gas analysis system contains a dust filter and water condenser, dry pyrolysis gas for analysis consists only of compounds that remain gaseous at 5°C. The composition was determined by a set of gas analysers for CO, CO_2, H_2, CH_4 and O_2 and by a micro gas chromatograph (µGC) for the same components plus N_2, Ar, C_2H_2, C_2H_4, C_2H_6, C_6H_6 and C_7H_8. As the O_2 and Ar signals of the µGC overlap, the analyser O_2 signal is used to obtain a correct Ar content.

The amount of other hydrocarbons in pyrolysis gas was determined during the endurance tests by capture from a measured amount of dry gas in washing bottles according to the tar protocol [2], followed by off-line analysis. As the procedure involves cooling to -20°C, water is captured too. Determination of the amount of water caught in the washing liquid allows calculation of the moisture content of the gas.

The composition of pyrolysis gas downstream of the tar protocol washing bottles was determined with the same analyser and µGC systems used upstream. Downstream of the tar sampling bottles, dry pyrolysis gas contains only components that remain gaseous at -20°C and do not dissolve in the washing liquid (isopropanol). Data from the gas analysers are stored at 1 minute intervals, data from the µGC about every 3 minutes.

Tar cracker fuel gas

The composition of dry tar cracker fuel gas was determined continuously by a second set of gas analysers and a second µGC. Again, data from the gas analysers are stored every minute and data from the µGC about every 3 minutes. The µGC used did not allow an accurate measurement of S-compounds. To get more detailed information on S-compounds, some gas samples were analysed off-line by a more sensitive GC analyser.

The soot content of the fuel gas was determined by collecting soot on a dust filter kept at 300ºC from a sample of about 50 liters of dry gas. Simultaneously, the moisture content of the gas was determined by weighing the amount of water collected in a washing bottle cooled to 5ºC.

The chlorine content of fuel gas was determined by capture of chlorine from a measured amount of gas in a washing liquid, followed by off-line chemical analysis of the washing liquid.

Fuel cell off-gas

The composition of dry off-gas from the SOFC stack was analysed with a third µGC. Unfortunately, measurements during the first two endurance tests were frequently interrupted by software problems. During the third endurance test, a GC was used to analyse the fuel cell off-gas and a continuous set of data was obtained.

3. Optimization of two-stage gasification process

3.1 Introduction

The aim of the two-stage gasification process is to convert the solid fuel into fuel gas with maximum thermal efficiency. The first step of the process is slow pyrolysis which produces pyrolysis gas and char. As for now, char is not used in the process. Hence, conditions should be chosen to minimize the energy content of char. The second step of the process is thermal tar cracking which breaks the pyrolysis gas down to a syngas mixture containing only CO, CO_2, H_2 and H_2O. The process conditions should be chosen to minimize the amount of soot, which will be formed as a by-product. As the composition of the pyrolysis gas is one of the variables to be considered, the process optimization has to be performed for both steps simultaneously. The parameters which can be varied are the fuel residence time in the pyrolysis unit, the steam to fuel ratio, the pyrolysis and tar cracking temperatures, the residence time in the tar cracker and the O_2 to fuel ratio in the tar cracker.

Although three different fuels have been used for the endurance tests, most optimization tests have been performed with beech wood chips. Further tests have been performed with Rofire pellets, which consist mainly of biomass but also contain a considerable amount of plastics. The third fuel is an industrial waste product, derived mainly from biological raw materials. Pictures of the three fuels are shown in Figure 3.1. The measured composition and heating values of the three fuels are given in Table 3.1. From the composition of the Rofire pellets we estimate they contain 30% to 35% plastics.

Figure 3.1 *Test fuels, from left to right: beech wood chips, Rofire pellets and industrial waste.*

Table 3.1 *Composition and heating values of fuels used in tests.*

		Wood	*Rofire*	*Industrial Waste*
Moisture	% ar	9.9	1.6	2.1
Ash	% dry	1.0	6.4	19.1 (30)[1]
C	% dry	47.2	58.6	45.2
H	% dry	5.8	8.1	6.0
O	% dry	45.9	28.2	29.3
N	% dry	0.1	0.2	<0.1
S	% dry	0.02	0.1	
Cl	% dry	0.005	1.9	0.009
HHV	MJ/kg dry	19.2	26.0	19.1
LHV	MJ/kg ar	15.9	23.7	17.3

[1] Lower value obtained at 815°C, higher value at 550°C.

3.2 Optimization tests

The main issue of the optimization tests was to find operating conditions which yield syngas with sufficiently low soot concentration to allow SOFC test runs of 100 hours. As the SOFC stack is protected against soot by the set of hot gas filters shown in Figure 2.3, the aim is in fact to reduce the soot concentration to an extent which allows a reasonable stand time of a filter.

In the previous project, a substantial flow of inert gas was used to flush pyrolysis gas from the pyrolysis unit to the tar cracker. In the present project steam was used instead. Steam acts as an agent to suppress soot formation and might react with char to yield additional syngas. The inert gas flow was reduced to the minimum required to prevent steam and pyrolysis gas from flowing to the fuel bunker. The temperature in the pyrolysis unit was kept at 550°C. A few tests performed at 500°C showed no significant differences.

The tar cracker is heated by combustion of CH_4 with O_2. If excess O_2 is supplied, some of the pyrolysis gas is burnt too. Although some heat is required for cracking of hydrocarbons in the pyrolysis gas, most of the heat is supplied to compensate heat loss to the surroundings. In an installation of commercial size, heat loss is relatively less important. The process heat can then be supplied by combustion of some of the pyrolysis gas.

The syngas produced by cracking of the pyrolysis gas is diluted by the combustion products of the CH_4/O_2 flame. In order to obtain a mixture more akin to the syngas which would be obtained in a commercial installation, some tests have been performed with substoichiometric conditions in the CH_4/O_2 flame, i.e. with 83% of the O_2 amount for complete combustion.

The temperature in the tar cracker was kept at a minimum of about 1200°C near the pyrolysis gas entrance. At lower temperatures the risk of soot formation increases. The temperature in the tar cracker drops gradually to about 800°C at the gas exit. The exit temperature increases with time of operation. Because of the high thermal inertia of the tar cracker, it takes several days before truly stable conditions are reached. The exit temperature also rises with total gas flow, as the heat loss becomes relatively less important. The gas residence time in the tar cracker varies from about 1.5 seconds at high temperature and flow to just over 2 seconds at lower flow and temperature.

3.2.1 Beech wood chips

Tests with beech wood chips have been performed with feed rates of 3 to 5 kg/hr. Results are shown in Figure 3.2 and Figure 3.3. The soot concentration decreases when steam is added, but above 0.3 kg steam per kg wood there is little further improvement. There is also a correlation between the soot concentration and the tar cracker temperature. Higher soot concentrations are observed at lower temperatures. Substoichiometric operation of the CH_4/O_2 flame that heats the tar cracker has no effect on the soot concentration.

A soot concentration below 0.5 g/Nm^3 can be obtained with a steam to fuel ratio of 0.3 or more. If 1 g/Nm^3 can be accepted, the steam to fuel ratio can even be lowered to about 0.15.

3.2.2 Rofire pellets

Tests with Rofire pellets have been performed with feed rates of 1.5 to 3 kg/hr. Results are shown in Figure 3.4 and Figure 3.5. Rofire pellets produce significantly more soot than beech wood, probably because of the presence of plastics. Again, the soot concentration decreases with steam addition. However, about 1.5 kg steam per kg Rofire is needed to reach a level of 1 g/Nm^3. There seems to be little correlation between the soot concentration and the tar cracker temperature. As for the case of wood, substoichiometric operation of the CH_4/O_2 flame that heats the tar cracker has no effect on the soot concentration.

A soot concentration below 1 g/Nm^3 requires a steam to fuel ratio of 1.5 or more. If 3 g/Nm^3 can be accepted, the steam to fuel ratio can be lowered to about 0.75.

3.2.3 Industrial waste

As there was only a limited amount of the industrial waste available for tests, and as it was mainly of biological origin, we have not performed optimization tests but directly chosen a steam to fuel ratio expected to limit soot production. Results of that approach are given in the next chapter.

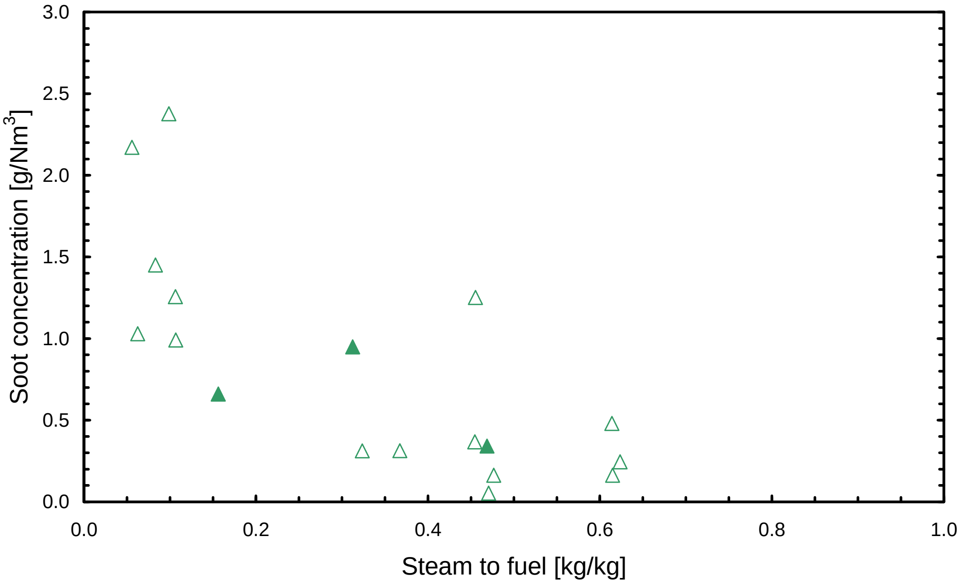

Figure 3.2 *Effect of steam to fuel ratio on soot concentration in syngas produced by cracking of pyrolysis gas from beech wood. Closed symbols represent substoichiometric conditions.*

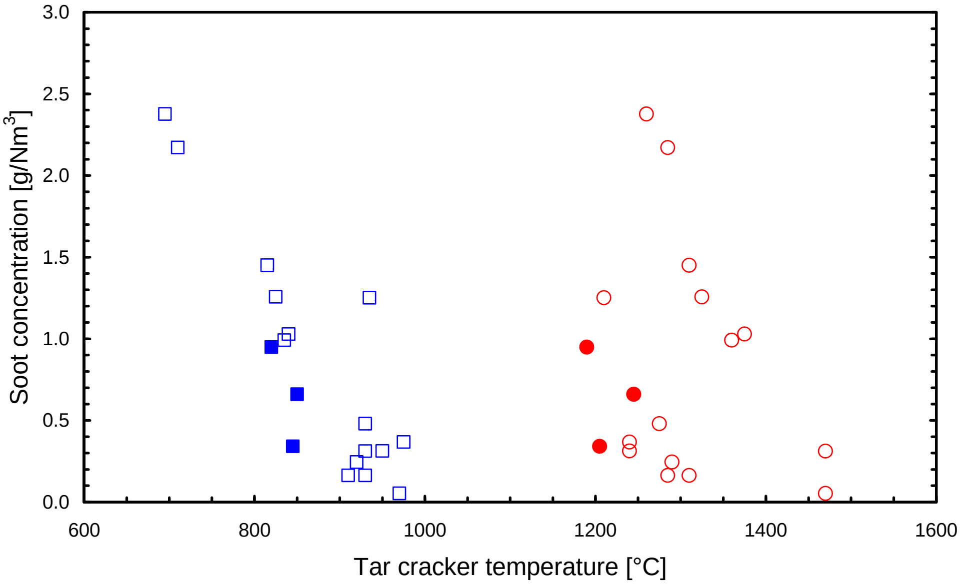

Figure 3.3 *Effect of temperature in tar cracker at entrance (circles) and exit (squares) on soot concentration in syngas produced by cracking of pyrolysis gas from beech wood. Closed symbols represent substoichiometric conditions.*

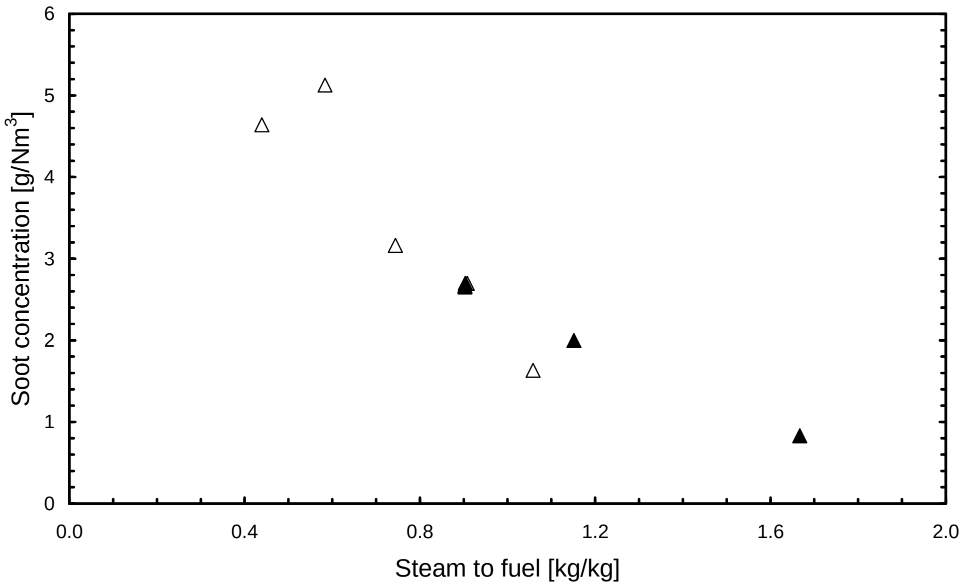

Figure 3.4 *Effect of steam to fuel ratio and tar cracker temperature on soot concentration in syngas produced by cracking of pyrolysis gas from Rofire pellets. Closed symbols represent substoichiometric conditions.*

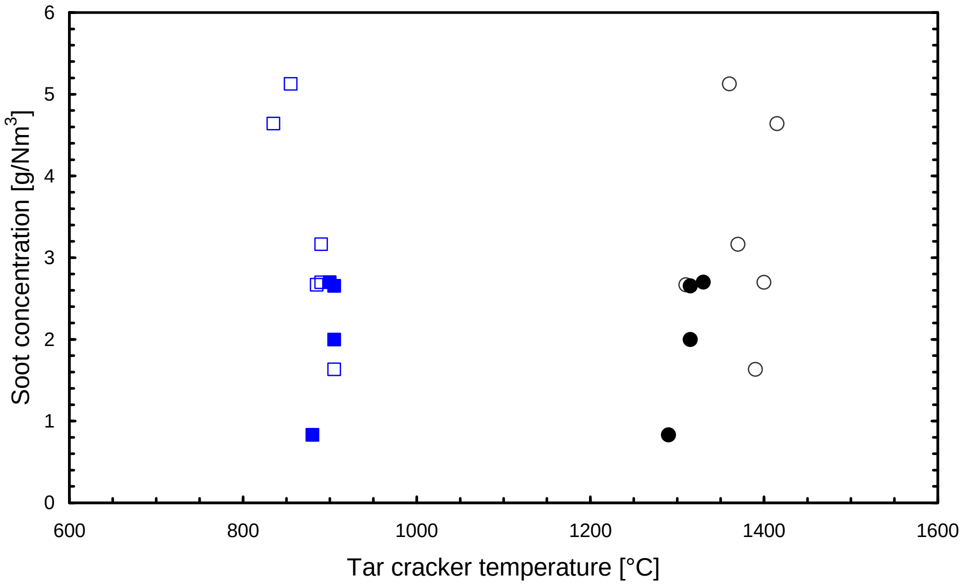

Figure 3.5 *Effect of temperature in tar cracker at entrance (circles) and exit (squares) on soot concentration in syngas produced by cracking of pyrolysis gas from Rofire pellets. Closed symbols represent substoichiometric conditions.*

4. Endurance tests

Endurance tests have been performed with the three fuels described in chapter 3. The SOFC stack was operated at a constant current of 10 A. The syngas flow was set at a value which kept the voltage of each set of three cells at or above 2.1 V. At lower voltages, there is an increasing risk of damage to the cells. Table 4.1 and Table 4.2 summarize the conditions and results of the endurance tests. In between tests, the SOFC stack was operated with a gas mixture of 40% H_2 and 60% N_2. The stack performance remained stable with negligible degradation during the first two tests. The third test did show signs of deteriorating stack performance, but that test was performed after more than 3000 operating hours. The tests are described in detail in the following sections.

Table 4.1 *Summary of conditions and results of endurance tests with three different fuels.*

		Wood	*Rofire*	*Industrial Waste*
Fuel feed rate	kg/hr	3.5	1.65	3.1
	kW (HHV)	16.8	11.9	16.1
Steam flow	kg/hr	1.0	2.3	1.0
Char production	kg/hr	0.74	0.34	1.32
	kW (HHV)	6.7	2.1	3.6
Test duration	hrs	101	99	54
SOFC syngas input [1]	kW (HHV)	0.726	0.717	0.736
SOFC off-gas	kW (HHV)	0.317	0.302	0.357
SOFC fuel utilization [1]	%	56	58	52
SOFC power output	kWe	0.225	0.219	0.215

[1] Comparison with results for H_2/N_2 and $H_2/CO/CO_2$ show, that the calculated syngas input and off-gas output are probably too low and the fuel utilization close to 53% in all three tests.

Table 4.2 *Composition and gross heating values of char from fuels used in tests.*

		Beech wood char	*Rofire char*	*Industrial waste char*
Ash	% dry	5.1	30.7	44.5
C	% dry	80.0	56.0	33.6
H	% dry	2.7	2.4	1.1
O	% dry	7.2	8.8	19.7
N	% dry	0.15		
S	% dry	0.04	0.12	
Cl	% dry	0.03	3.7	
HHV	MJ/kg dry	32.1	23	9.9

4.1 Beech wood test

General description

The first endurance test was performed with beech wood chips and a steam to fuel ratio of 0.3 (i.e. 1 kg/hr steam to 3.5 kg/hr wood). The fuel feed rate corresponds to 16.8 kW gross and 15.5 kW net thermal input.

The test started at the 12th of November 2007 at 8:30. The SOFC stack obtained cleaned tar cracker syngas from 12:00. The gas supply ended at the 16th of November at 13:15. The test was interrupted three times for 1 to 2 hours. In two cases, the feeding screw bent, when the wood chips stuck in the feeding tube. In both cases, the screw had to be replaced by a new one. In order to prevent further feeding problems, the inert gas flow along the screw was increased and, between the first and second endurance tests, the steam entrance point was moved to a position further away from the fuel feeding point.

The third interruption was necessary because of the increasing pressure drop over the pipe connecting the pyrolysis unit with the tar cracker. By addition of a small amount of air to the pyrolysis gas, we succeeded in burning away the soot deposited on the tube wall. During the interruptions, the SOFC stack was switched to operation on a similar synthetic gas mixture.

Three times over the 100 hours test period, accumulation of soot on the high-temperature filter downstream of the tar cracker (see Figure 2.3) caused such an increase of pressure drop that it was necessary to switch from one filter to the other. Gradually, the gas pipe from the tar cracker to the filter got clogged too. That problem could be solved easily by a burst of inert gas to blow the soot back into the tar cracker.

The fuel bunker was filled with 15 kg of wood at the start. The bunker was refilled when about 30% of the fuel had been fed to the pyrolysis unit. At the end of the test, and when the screw had to be replaced, the bunker was emptied and its contents weighed. Based on the weight measurements, the average fuel feed rate was 3.5 kg/hr with the first and third feeding screws and 3.2 kg/hr with the second screw.

Char and ash were collected in vessels which were replaced about every two hours. Used vessels were put in a water bath to cool, and weighed when cool. The average amount of char collected equalled 0.74 kg/hr, i.e. 21% of the fuel input. The char contains 5% ash, 80% C, 2.7% H and 7.2% O. It has a gross heating value of 32 MJ/kg and a net heating value of 31.4 MJ/kg. The char production corresponds to 6.7 kW, which means char represents 40% of the gross heating value and 42% of the net heating value of the wood fuel.

Pyrolysis gas

The composition of pyrolysis gas has been determined in two periods of about half an hour when samples were taken according to the tar protocol. Due to air leaking into the sampling system at the low pressure needed to pump tar-laden gas through the system, the raw data showed 1% to 1.6% O_2 in pyrolysis gas. Table 4.3 gives the result for dry gas after correction for air leakage. As the gas first passed through washing bottles filled with isopropanol at -20°C, heavy hydrocarbons were not detected.

Analysis of the contents of the washing bottles indicated that the water content of the pyrolysis gas was 77%. The gas contained 730 g/Nm^3 tar in total. Table 4.4 shows the components contributing at least 1 g/Nm^3 dry gas. About 50% of the total was not identified.

Table 4.3 *Composition of wood pyrolysis gas (dry basis, heavy hydrocarbons removed)*[1].

Component	*Unit*	*Concentration*	*Component*	*Unit*	*Concentration*
H_2	%	7.2	C_2H_4	%	1.5
CO	%	33.1	C_2H_6	%	1.3
CO_2	%	25.3	C_2H_2	%	0.02
CH_4	%	12.3	H_2S	ppm	68
N_2	%	16.6	COS	ppm	9
O_2/Ar	%	0.4			

[1] Data corrected for air leakage into gas sample due to pressure drop over gas sampling and conditioning system.

Table 4.4 *Composition of tar in beech wood pyrolysis gas (dry basis, g/Nm^3). Data not corrected for air leakage.*

Component	*Formula*	*Conc.*	*Component*	*Formula*	*Conc.*
Acetaldehyde	C_2H_4O	21	5-Methyl-2-furaldehyde	$C_6H_6O_2$	1.1
Propanal	C_3H_6O	9	Furan-2-methanol	$C_5H_6O_2$	4
Furan	C_4H_4O	5	2(5H)-furanon	$C_4H_4O_2$	1.3
Acetone	C_3H_6O	14	Phenol	C_6H_6O	4
Methylacetate	$C_3H_6O_2$	2.5	5-(Hydoxymethyl)-2-furaldehyde	$C_6H_6O_3$	2.5
Methanol	CH_4O	35	Pyrocatechol	$C_6H_6O_2$	19
2-Butenal	C_4H_6O	2.5	Hydroquinone	$C_6H_6O_2$	1.4
Hydroxyacetone	$C_3H_6O_2$	37	Levo-glucosan	$C_6H_{10}O_5$	24
1-Hydroxy-2-butanone	$C_4H_8O_2$	6			
Acetic acid	$C_3H_6O_2$	154	Unidentified		380
2-Furaldehyde	$C_5H_4O_2$	7	Total		730

Syngas

Pyrolysis gas is fed to the tar cracker, which is heated by combustion of 2.7 Nm^3/hr CH_4 with 4.5 Nm^3/hr O_2. The substoichiometric conditions have been chosen to simulate gas produced by a large system in which no CH_4 would be added but part of the pyrolysis gas would be burned. Figure 4.1 shows the measured temperatures in the tar cracker. Close to the entrance, the temperature was about 1300°C. The three peaks show the test interruptions, when there was no pyrolysis gas entering the tar cracker. Along the tar cracker, the temperature decreases by heat loss and endothermic reactions. Near the exit, the temperature has dropped to 920°C.

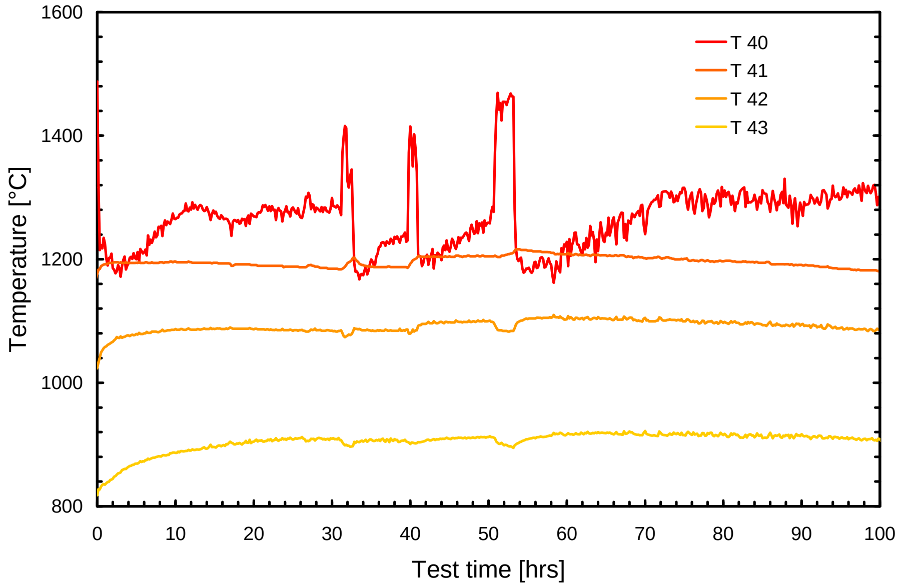

Figure 4.1 *Temperatures in tar cracker from entrance (T40) to exit (T43).*

Downstream of the tar cracker and upstream of the hot gas filter, the soot content of the syngas is about 0.6 g/Nm^3 wet gas, in line with results shown in Figure 3.2. The syngas contains 43% moisture. Figure 4.2 and Figure 4.3 show the composition of dry syngas downstream of the hot gas filter. The O_2 content is an artefact due to gas leaking into the analysis system. It peaks periodically when the gas analysis guard filter gets clogged by soot, resulting in high pressure drop over the filter. The analysis also shows that some CH_4 (about 0.5%) remains in the syngas. About halfway the test, the N_2 content of the syngas increases because of the higher flow applied to the feeding screw.

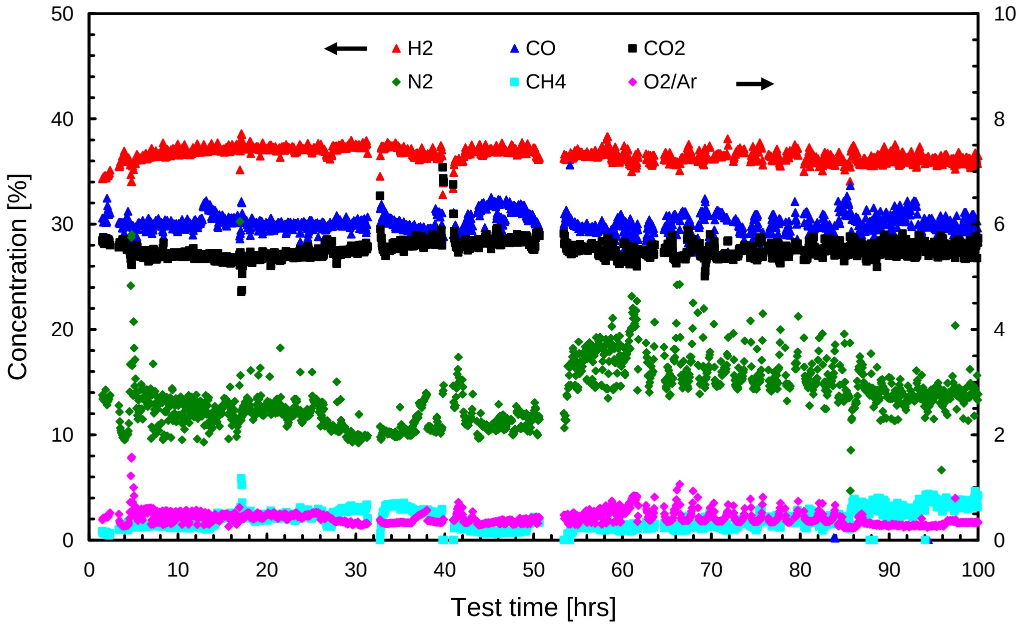

Figure 4.2 *Composition of dry syngas from beech downstream of tar cracker.*

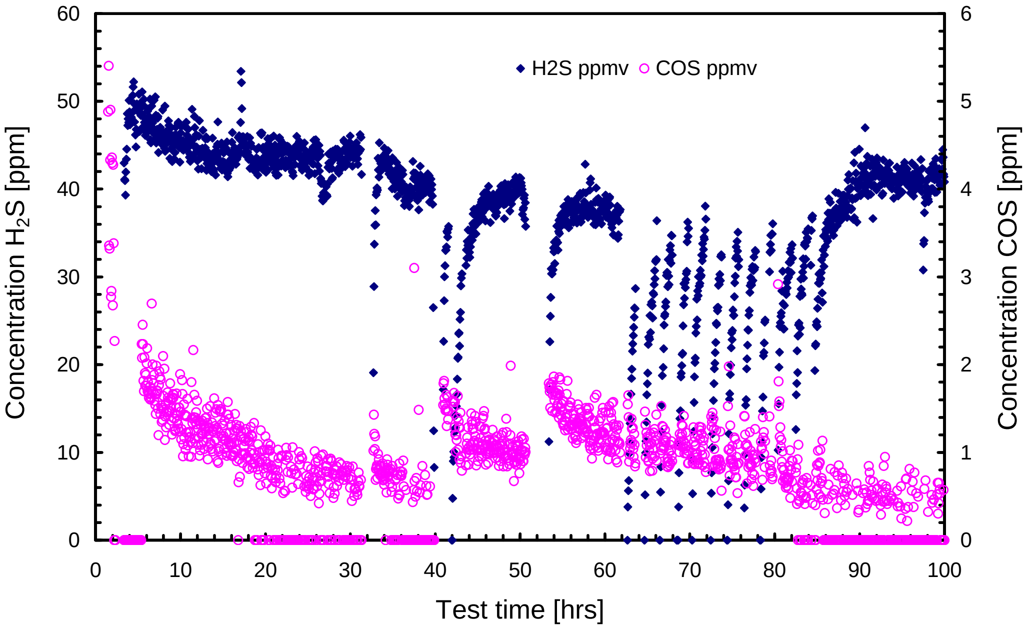

Figure 4.3 *Concentrations of H_2S and COS in dry syngas from beech downstream of tar cracker.*

Continuous measurements of the H_2S and COS concentrations downstream of the gas cleaning indicate complete removal of these components. Measurements with a more sensitive detector on gas samples taken once a day showed confirmed the results, except for the last day when 30 ppb of both H_2S and COS were detected. These measurements also showed an increase of the COS concentration upstream of the gas cleaning, from 3 ppm at the first day, 5 ppm at the second day, 10 ppm at the third day to 18 ppm at the fourth day. These values are significantly higher than those shown in Figure 4.3. Although the cause of the discrepancy is not certain, it is known that the sampling system can reduce the level of S components by adsorption. For that reason, results have to be considered with some caution, especially for raw gas containing high concentrations.

The Cl concentration in syngas upstream and downstream of the gas cleaning has also been determined dayly. To that end, 30 to 50 liters of syngas were passed through a washing bottle and the contents of the washing fluid analysed afterwards. According to these measurements, the syngas contained 1 to 5 mg/Nm^3 Cl upstream and 0.2 to 0.5 mg/Nm^3 Cl downstream the gas cleaning.

SOFC performance
The SOFC stack was tested with constant 10 A load, except for interruptions due to problems or regular measurements of the impedance or I-V curves. The average stack voltage was 22.5 V, which means the stack delivered 225 W electrical power. Figure 4.4 shows the measured stack voltage and voltages delivered by the 10 sets of three cells. At the start of the test, set 1 delivers a lower voltage than the other sets. After optimization of the operating conditions, aimed at a more homogeneous temperature distribution over the stack, the performance improves and becomes indistinguishable from the other ones.

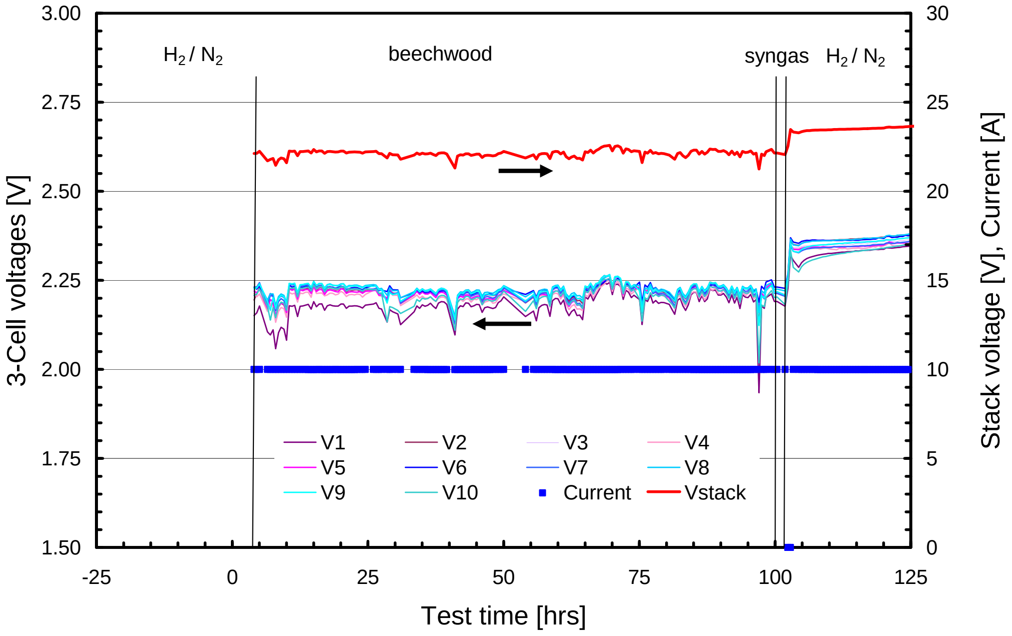

Figure 4.4 *SOFC performance during test with syngas from beech.*

Immediately after the test with beech syngas, the stack was fed with synthetic syngas of approximately the same composition (40% H_2, 30% CO and 30% CO_2 in dry gas, water content 40%). The stack performance did not change. When the stack was switched to a dry gas mixture of 40% H_2 and 60% N_2 the output rose quickly to 233 W, followed by a gradual increase to 237 W after 24 hours and 240 W after 72 hours. The lower performance on syngas is mainly caused by the presence of water and the lower H_2 partial pressure which results in a lower so-called Nernst voltage. The slow "recovery" of sets 1 and 10 on H_2/N_2 is probably due to a change in stack temperature.

Table 4.5 shows the average composition of dry syngas upstream and downstream the SOFC stack. As the syngas should not contain O_2, the O_2/Ar value should be accurate for Ar. However, air leakage into the gas sampling system is observed when samples are taken at less than atmospheric pressure (see e.g. Figure 4.2). If the Ar concentration in off-gas is correct and the C balance reliable, the true Ar concentration in dry input syngas can be calculated, as will be explained below. The Ar concentration thus obtained is given between brackets.

Based on the Ar content in the off-gas and the Ar flow added to the input gas, the dry off-gas flow would be 0.255 Nm^3/hr. From the C balance the dry gas input flow is calculated to be 0.328 Nm^3/hr. From that value a corrected Ar concentration of 4.2% is derived. From the H balance, the off-gas water content is calculated to be 53%. The dry syngas flow and composition correspond to a thermal input to the SOFC of 726 W (HHV) [1] or 664 W (LHV). The dry off-gas flow and composition correspond to 317 W (HHV) or 288 W (LHV). The fuel utilization by the SOFC is 56.5%. The SOFC electrical efficiency is 31.0% (HHV) or 33.9% (LHV). The remaining 25.3% (HHV) or 22.7% (LHV) are converted into heat.

For the synthetic syngas, the fuel input was equivalent to 0.846 kW (HHV) or 0.771 kW (LHV). The fuel utilization calculated from the C balance is 56.8%. A much lower value of 52.2% is calculated from the current drawn and the fuel input. The electrical efficiency was 26.4% (HHV) or 28.9% (LHV). For the H_2/N_2 mixture, the fuel input corresponded to 0.849 kW (HHV) or 0.719 kW (LHV) and the electrical efficiency (after recovery) 28.3% (HHV) or 33.4% (LHV).

Table 4.5 *Average composition of dry syngas from beech upstream and downstream of SOFC stack. Water content of syngas 41% upstream and 53% downstream of SOFC.*[2]

		SOFC in	SOFC out
H_2	%	34.2	20.8
CO	%	27.5	14.4
CO_2	%	27.1	56.3
CH_4	%	0.36	0.04
N_2	%	3.4	4.3
O_2/Ar (Ar)	%	4.7 (4.2)	5.4

Discussion

If the measured Ar concentration in the input gas has to be corrected for air leakage into the gas analysis system, the N_2 concentration should be corrected too. Here, it would ruin the N_2 balance. The discrepancy can be explained if air leaked directly into the syngas. The N_2 from leaked air could not be discerned from N_2 added in an earlier stage. The O_2 would affect the Ar signal in the input gas but not in the off-gas. It would just burn some of the syngas in the SOFC and reduce the SOFC electrical efficiency. The second explanation fits the observed difference in the Ar and N_2 balances.

Another explanation for the observed discrepancy is simply the accuracy of the two µGC's. The instruments are calibrated with a gas mixture similar to the gas to be analysed. If the actual gas composition differs, the calibration factors can be wrong. As that could be the case for either of the two µGC's, the calculated syngas flow to the SOFC stack could be wrong. The results obtained with synthetic syngas suggest that the beech syngas flow may have been higher and the electrical efficiency lower than calculated.

4.2 Rofire test

General description

The second endurance test was performed with Rofire pellets and a steam to fuel ratio of 1.4 (i.e. 2.3 kg/hr steam to 1.65 kg/hr pellets). The fuel feed rate corresponds to 11.9 kW gross and 10.9 kW net thermal input.

The test started at the 19th of November 2007 at 10:30. The SOFC stack obtained cleaned tar cracker syngas from 11:30. The gas supply ended at the 23th of November at 13:30. There were no major interruptions to the test. There were only a few short breaks to switch gas filters or remove a deposit in the connection between the pyrolysis unit and tar cracker.

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[1] The HHV value of the gas refers to dry gas, i.e. the heat of condensation of the water content has not been taken into account.

[2] The water content of syngas upstream of the SOFC is lower than that of syngas downstream of the tar cracker because of dilution by the addition of Ar.

The fuel bunker was filled with 15 kg of Rofire pellets at the start. It was intended to refill the bunker when about 40% of the fuel had been fed to the pyrolysis unit. After one day, the interval between refills was halved to obtain more stable conditions.

Char and ash were collected in vessels which were replaced about every hour and a half. Used vessels were put in a water bath to cool, and weighed when cool. The average amount of char collected equalled 0.34 kg/hr, i.e. 21% of the fuel input. As the Rofire char was sticky, the valves through which char passes from the pyrolysis unit to the char vessel had to be cleaned regularly. The char contains 31% ash, 56% C, 2.4% H and 9% O. It has a gross heating value of 22.5 MJ/kg and a net heating value of 22.0 MJ/kg. The char production corresponds to 2.1 kW, which means char represents 18% of the gross heating value and 19% of the net heating value of the Rofire fuel. The char contains 3.7% Cl and 0.12% S, which means Cl is concentrated by a factor 2 and S by 20%.

Pyrolysis gas

The composition of pyrolysis gas has been determined in two periods of about half an hour when samples were taken according to the tar protocol. Due to air leaking into the sampling system at the low pressure needed to pump tar-laden gas through the system, the raw data showed 3% to 4% O_2 in pyrolysis gas. Table 4.6 gives the result for dry gas after correction for air leakage. As the gas first passed through washing bottles filled with isopropanol at -20°C, heavy hydrocarbons were not detected.

Table 4.6 *Composition of Rofire pyrolysis gas (dry basis, heavy hydrocarbons removed)*[1].

Component	*Unit*	*Concentration*	*Component*	*Unit*	*Concentration*
H_2	%	7.7	C_2H_4	%	3.0
CO	%	11.3	C_2H_6	%	2.2
CO_2	%	18.8	C_2H_2	%	0.02
CH_4	%	6.4	H_2S	ppm	750
N_2	%	46.6	COS	ppm	18
O_2/Ar	%	0.3			

[1] Data corrected for air leakage into gas sample due to pressure drop over gas sampling and conditioning system.

The N_2 content of Rofire pyrolysis gas is substantially higher than that of beech pyrolysis gas (compare Table 4.3 and Table 4.6). It is the result of the lower fuel feed rate and the higher N_2 flow to the feed bunker, which was increased halfway the first endurance test. Rofire pyrolysis gas contains much less CO and significantly more C_2H_4 and C_2H_6. The concentration of H_2S is much higher. The Cl content has not been determined, but must have been in the order of 1%.

Analysis of the contents of the tar protocol washing bottles indicated that the water content of the pyrolysis gas was 88%. The washing bottles contained a large amount of styrene (C_8H_8), corresponding to an estimated 500 g/Nm^3 dry gas. Styrene is produced by decomposition of plastics in the Rofire pellets. Analysis of the tar samples showed dry Rofire pyrolysis gas contained 580 g/Nm^3, nearly 70% of which from unidentified components. Table 4.7 shows the components contributing at least 1 g/Nm^3. Rofire tar contains relatively less methanol and acetic acid and more aldehydes than beech wood tar.

Table 4.7 *Composition of tar in Rofire pyrolysis gas (dry basis, g/Nm^3). Data not corrected for air leakage.*

Component	*Formula*	*Conc.*	*Component*	*Formula*	*Conc.*
Acetaldehyde	C_2H_4O	34	5-Methyl-2-furaldehyde	$C_6H_6O_2$	4
Propanal	C_3H_6O	6	Furan-2-methanol	$C_5H_6O_2$	-
Furan	C_4H_4O	5	2(5H)-furanon	$C_4H_4O_2$	2.0
Acetone	C_3H_6O	14	Phenol	C_6H_6O	4
Methylacetate	$C_3H_6O_2$	-	5-(Hydoxymethyl)-2-furaldehyde	$C_6H_6O_3$	4
Methanol	CH_4O	6	Pyrocatechol	$C_6H_6O_2$	5
2-Butenal	C_4H_6O	2.7	Hydroquinone	$C_6H_6O_2$	-
Hydroxyacetone	$C_3H_6O_2$	24	Levo-glucosan	$C_6H_{10}O_5$	18
1-Hydroxy-2-butanone	$C_4H_8O_2$	2.6			
Acetic acid	$C_3H_6O_2$	41	Unidentified		395
2-Furaldehyde	$C_5H_4O_2$	12	Total		580

Syngas

Pyrolysis gas is fed to the tar cracker, which is heated by combustion of 2.7 Nm^3/hr CH_4 with 4.5 Nm^3/hr O_2. The substoichiometric conditions have been chosen to simulate gas produced by a large system in which no CH_4 would be added but part of the pyrolysis gas would be burned. Figure 4.5 shows the measured temperatures in the tar cracker. Close to the entrance, the temperature was about 1350°C. The dip near 70 hours occurred when the connection between the pyrolysis unit and the tar cracker was cleaned by a gas pulse. The dip near 37 hours coincides with cleaning of the gas analysis, which may also have caused a temporary disturbance in the gas flow. Along the tar cracker, the temperature decreases by heat loss and endothermic reactions. Near the exit, the temperature has dropped to about 900°C. Thermocouple T 41 failed shortly before the end of the test.

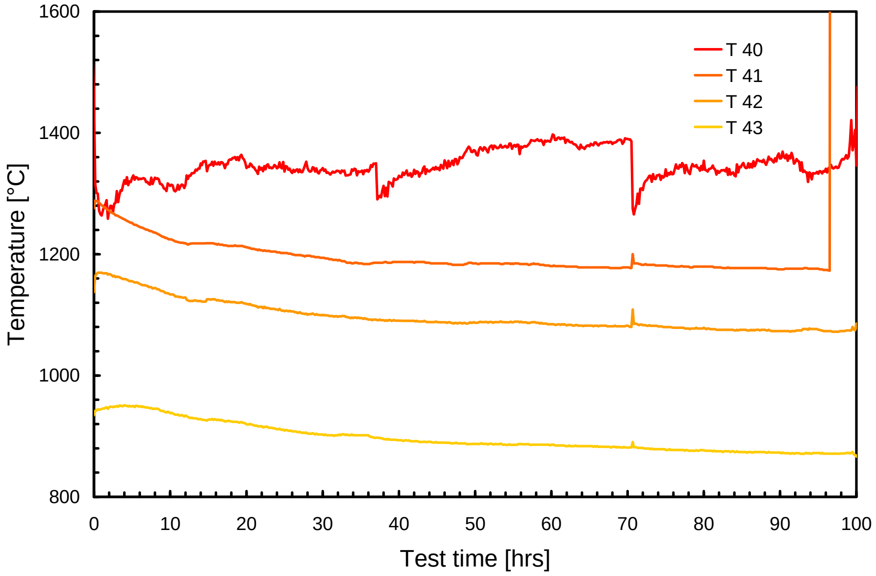

Figure 4.5 *Temperatures in tar cracker from entrance (T40) to exit (T43).*

Downstream of the tar cracker and upstream of the hot gas filter, the soot content of the syngas is about 1.6 g/Nm^3 wet gas, in line with results shown in Figure 3.4. The syngas contains 46% moisture. Figure 4.6 and Figure 4.7 show the composition of dry syngas downstream of the hot gas filter. The Rofire syngas contains a little more H_2 and less CO than the beech syngas. The low N_2 content between 12 and 15 hours and the high N_2 content between 32 and 36 hours are due to operator errors. The H_2S and COS results are shown for the first 48 hours only. Later data with a different gas sampling system were unreliable.

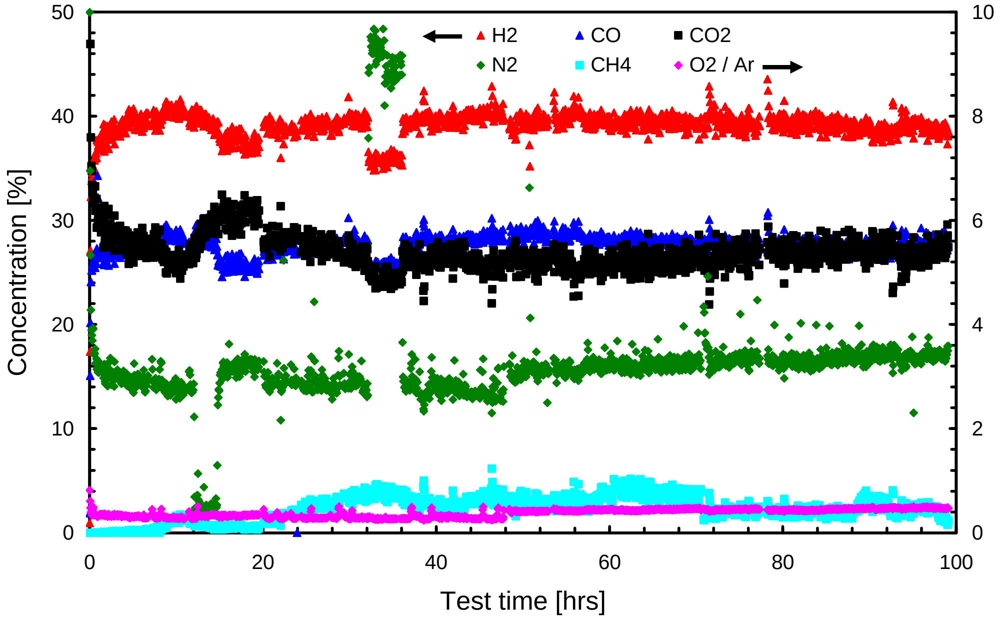

Figure 4.6 *Composition of dry syngas from Rofire downstream of tar cracker.*

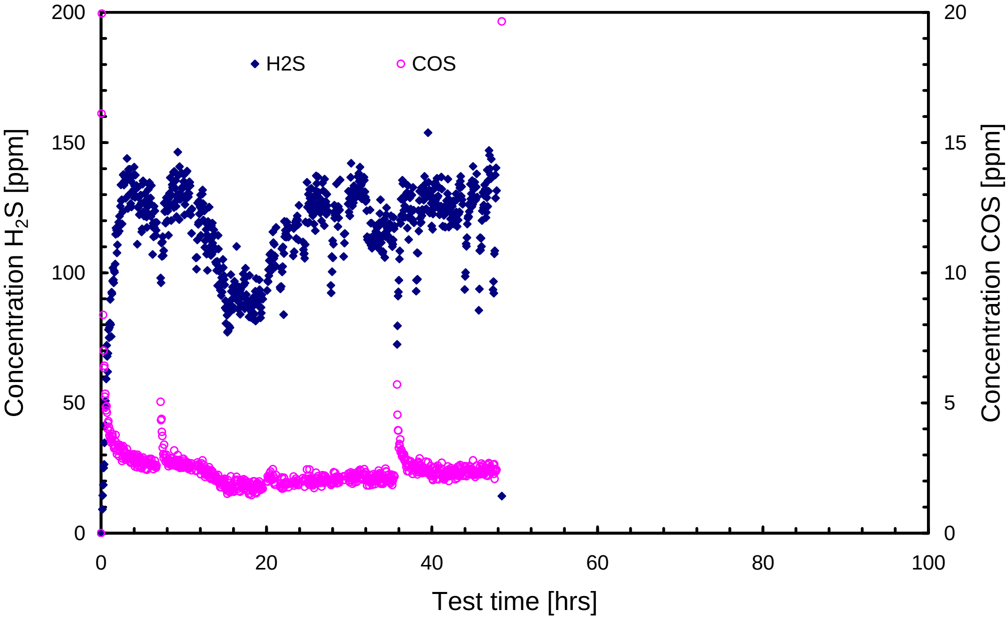

Figure 4.7 *Concentrations of H_2S and COS in dry syngas from Rofire downstream of tar cracker.*

Continuous measurements of the H_2S and COS concentrations downstream of the gas cleaning indicate complete removal of these components. Measurements with a more sensitive detector on gas samples taken once a day confirmed the results (to the 50 ppb detection limit). Upstream of the gas cleaning, COS results were out of range (i.e. above 10 ppm) at the first and second day, 5.7 ppm at the third day and 7.5 ppm at the fourth day. These values are higher than those shown in Figure 4.7, but the discrepancy is smaller than for the previous test.

The Cl concentration in syngas upstream and downstream of the gas cleaning has also been determined dayly. To that end, 30 to 50 liters of syngas were passed through a washing bottle and the contents of the washing fluid analysed afterwards. According to these measurements, the syngas contained 1100 to 1300 mg/Nm^3 Cl upstream and 0.5 to 0.6 mg/Nm^3 Cl downstream the gas cleaning.

SOFC performance

The SOFC stack was tested with constant 10 A load, except for interruptions due to problems or regular measurements of the impedance or I-V curves. The average stack voltage was 21.9 V, which means the stack delivered 219 W electrical power. Figure 4.8 shows the measured stack voltage and voltages delivered by the 10 sets of three cells. Near the end of the test, the gas flow was reduced to study the behaviour at higher fuel utilization. The stack voltage decreased to 20 V and the difference between the weakest and strongest sets of three cells increased from 0.05 V to 0.25 V.

After the test with Rofire syngas, the stack was fed with a gas mixture of 40% H_2 and 60% N_2 and the output rose quickly to 235 W, followed by a gradual increase to 238 W after 24 hours and 243 W after 72 hours. This is practically the same behaviour observed after the first test. In fact, the performance is even a little bit better.

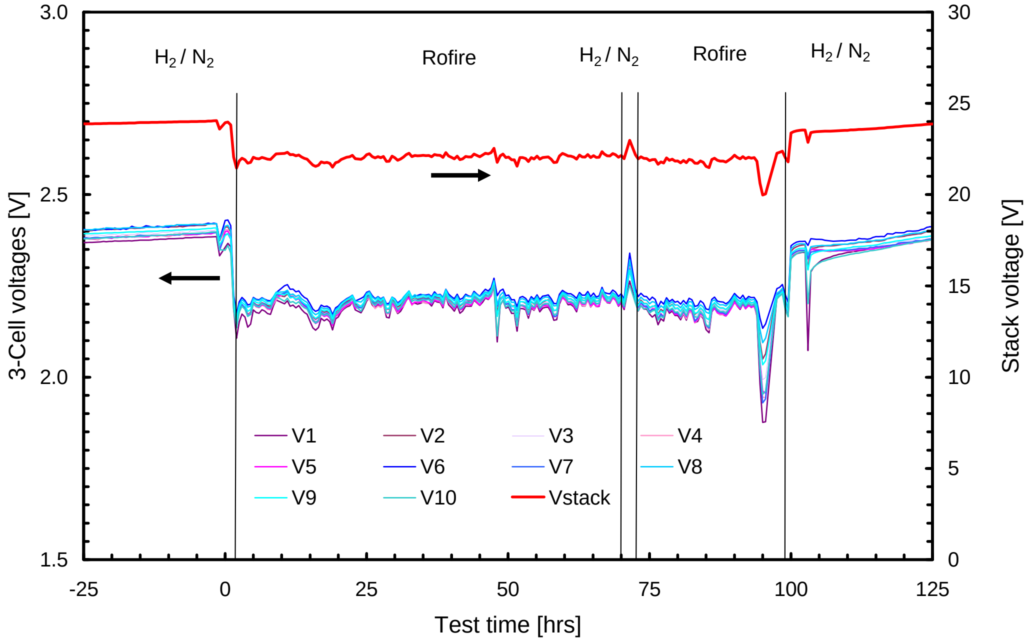

Figure 4.8 *SOFC performance during test with syngas from Rofire.*

Table 4.8 *Average composition of dry syngas from Rofire upstream and downstream of SOFC stack. Water content of syngas 44% upstream and 56% downstream of SOFC.*

		SOFC in	SOFC out
H_2	%	36.3	22.3
CO	%	25.6	12.6
CO_2	%	26.3	56.1
CH_4	%	0.41	0.01
N_2	%	3.6	3.6
O_2/Ar (Ar)	%	5.1 (4.3)	5.6

Table 4.8 shows the average composition of syngas upstream and downstream the SOFC stack. The Ar value between brackets was derived by the method explained in Section 4.1 to make the Ar balance confirm to the C balance. Here, the N_2 concentration in the syngas should also be corrected to make the N balance agree. Based on the Ar content in the off-gas and the Ar flow added to the input gas, the dry off-gas flow would be 0.245 Nm^3/hr. From the C balance the dry gas input flow is calculated to be 0.322 Nm^3/hr. From the H balance, the off-gas water content is calculated to be 56%.

The dry syngas flow and composition correspond to a thermal input to the SOFC of 717 W (HHV) or 652 W (LHV). The dry off-gas flow and composition correspond to 302 W (HHV) or 272 W (LHV). The fuel utilization by the SOFC is 58.2%. The SOFC electrical efficiency is 30.5% (HHV) or 33.6% (LHV). The remaining 27.3% (HHV) or 24.7% (LHV) are converted into heat.

Unfortunately, problems with the off-gas analysis made it difficult to determine the fuel utilization during the short period at the end of the test. The Ar concentration in the syngas suggests the flow was reduced to 75%, in which case the utilization would have been 77%.

Discussion
As mentioned in the discussion at the end of Section 4.1, the calculated syngas flow may be too low and the electrical efficiency too high.

Although the test was very successful, post-test inspection of the gas cleaning system showed the use of Cl rich gas is not without risk. Some of the stainless steel gas tubes had been corroded all through the 1 mm wall thickness. The corrosion attacked preferentially at squeeze joints where the material was deformed. Although the tubes should have been kept at 200°C by trace heating, the local temperature may have been low enough to allow Cl condensation. No damage was observed near the hot gas filters, which operated at 300°C, or downstream of the Cl removal. The damaged parts were replaced before the third endurance test.

4.3 Industrial waste test

General description
The third endurance test was performed with industrial waste and a steam to fuel ratio of 0.32 (i.e. 1.0 kg/hr steam to 3.1 kg/hr waste). The fuel feed rate corresponds to 16.1 kW gross and 15.0 kW net thermal input.

The test started at the 18th of March 2008 at 8:00. The SOFC stack obtained cleaned tar cracker syngas from 10:30. The gas supply ended at the 20th of March at 14:00. There were one interruption of one and a half hours and a few short breaks to switch gas filters or remove a deposit in the connection between the pyrolysis unit and tar cracker.
The fuel bunker was filled with 8 kg fuel at the start and refilled with 4 kg about every hour and a quarter. The fuel level was kept low to prevent sticking of the fuel.

Char and ash were collected in vessels which were replaced about every hour and a half. Used vessels were put in a water bath to cool, and weighed when cool. The average amount of char collected equalled 1.32 kg/hr, i.e. 42% of the fuel input. The char contains 44.5% ash, 33.6% C, 1.1% H and 19.7% O. It has a gross heating value of 9.9 MJ/kg and a net heating value of 9.7 MJ/kg. The char production corresponds to 3.6 kW, which means char represents 23% of the gross heating value and 24% of the net heating value of the fuel.

Pyrolysis gas
The composition of pyrolysis gas has been determined in a period of about twenty minutes when samples were taken according to the tar protocol. Table 4.9 shows the results. No correction was necessary for air leaking into the analysis system. The composition has also been determined for two periods of similar duration in which tar and moisture were removed only by a cooled filter and not by the tar protocol washing bottles. Results of those measurements are shown between brackets in Table 4.9. The industrial waste pyrolysis gas contains more H_2, C_2H_4 and C_2H_6 than beech pyrolysis gas and less CO (c.f. Table 4.3). The higher N_2 content is the result of the lower net fuel input and somewhat higher N_2 flow to the feed bunker.

Analysis of the contents of the washing bottles indicated that the water content of the pyrolysis gas was 74%. The gas contained 816 g/Nm^3 tar in total. Table 4.10 shows the components contributing at least 1 g/Nm^3 dry gas. About 50% of the total was not identified.

Table 4.9 *Composition of dry industrial waste pyrolysis gas downstream of the tar protocol washing bottles. Values between brackets were obtained upstream of the washing bottles.*

Component	*Unit*	*Concentration*	*Component*	*Unit*	*Concentration*
H_2	%	14.6 (14.1)	C_2H_4	%	3.4 (5.1)
CO	%	10.2 (11.3)	C_2H_6	%	3.1 (3.0)
CO_2	%	22.6 (20.9)	C_2H_2	%	0.01 (0.03)
CH_4	%	11.6 (11.8)	H_2S	ppm	30 (50)
N_2	%	31.8 (30.2)	COS	ppm	2.3 (2.6)
O_2/Ar	%	0.3 (0.2)	C_6H_6	ppm	--- (1800)
			C_7H_8	ppm	--- (800)

Table 4.10 *Composition of tar in industrial waste pyrolysis gas (dry basis, g/Nm^3).*

Component	*Formula*	*Conc.*	*Component*	*Formula*	*Conc.*
Acetaldehyde	C_2H_4O	35	5-Methyl-2-furaldehyde	$C_6H_6O_2$	1.0
Propanal	C_3H_6O	17	Furan-2-methanol	$C_5H_6O_2$	1.3
Furan	C_4H_4O	2.4	2(5H)-furanon	$C_4H_4O_2$	0.8
Acetone	C_3H_6O	22	Phenol	C_6H_6O	3
Methylacetate	$C_3H_6O_2$	1.2	5-(Hydoxymethyl)-2-furaldehyde	$C_6H_6O_3$	2.0
Methanol	CH_4O	16	Pyrocatechol	$C_6H_6O_2$	5
Hydroxyacetone	$C_3H_6O_2$	16	Hydroquinone	$C_6H_6O_2$	-
Hydroxyacetaldehyde	$C_2H_4O_2$	9	Levo-glucosan	$C_6H_{10}O_5$	4
1-Hydroxy-2-butanone	$C_4H_8O_2$	3.4			
Acetic acid	$C_3H_6O_2$	46	Unidentified		614
2-Furaldehyde	$C_5H_4O_2$	3	Total		816

Syngas

Pyrolysis gas is fed to the tar cracker, which is heated by combustion of 2.4 Nm^3/hr CH_4 with 4.1 Nm^3/hr O_2. In the course of the test, these flows were reduced in a few steps to 2.1 and 3.8 Nm^3/hr. The substoichiometric conditions have been chosen to simulate gas produced by a large system in which no CH_4 would be added but part of the pyrolysis gas would be burned. Figure 4.9 shows the measured temperatures in the tar cracker. Close to the entrance, the temperature was about 1400°C for most of the test. Along the tar cracker, the temperature decreases by heat loss and endothermic reactions. Because of the lower gas flow, when compared with the other tests, heat loss has a larger effect. At the third and fourth thermocouples, the temperature is 100°C to 150°C lower than in the previous tests.

Downstream of the tar cracker and upstream of the hot gas filter, the soot content of the syngas is 4 to 5 g/Nm^3 wet gas. Apparently, the industrial waste fuel behaves much like Rofire (see Figure 3.4), even though it is derived from biological raw materials. The relatively low temperature in the second part of the tar cracker may also be to blame.

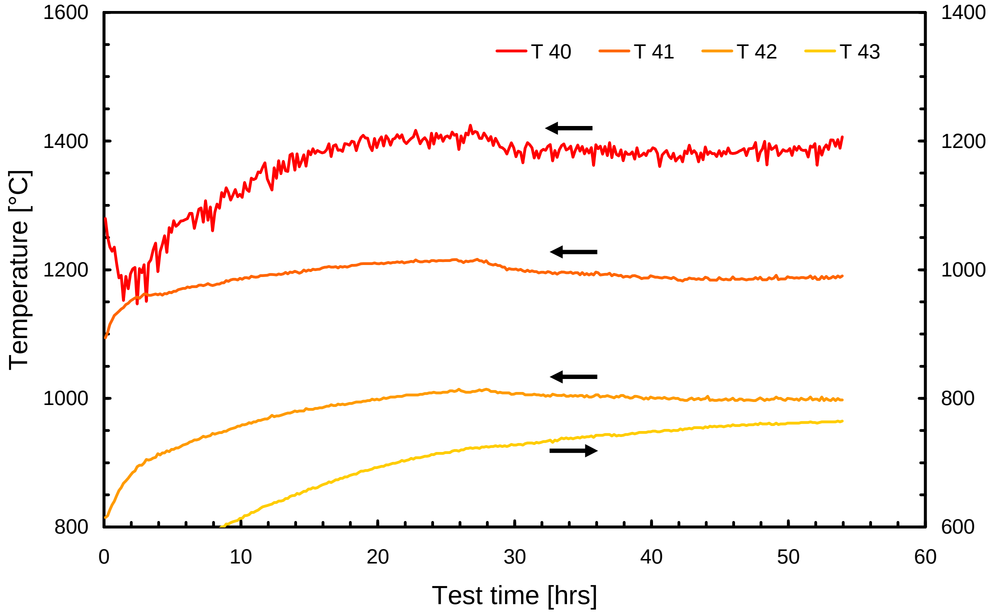

Figure 4.9 *Temperatures in tar cracker from entrance (T40) to exit (T43).*

The syngas contains 38% moisture. Figure 4.10 and Figure 4.11 show the composition of dry syngas downstream of the hot gas filter. It contains somewhat less H_2 and CO_2 than the beech syngas. The N_2 and O_2 content increase steadily, but drop at 41.5 hours when the filter upstream of the gas analysis was replaced. Gas analysis downstream of the gas cleaning shows a different pattern, with increasing N_2 and O_2 levels when the hot syngas filter got plugged. The syngas contains about 1% CH_4 and detectable levels of C_2H_4, C_2H_2 and C_6H_6. The presence of these components indicates the tar cracker temperature was too low. The H_2S and COS concentrations remained below the detection limits. Measurements with a more sensitive detector on gas samples taken once a day showed 4 ppm of both H_2S and COS. Downstream of the gas cleaning the levels were below the detection limit of 50 ppb.

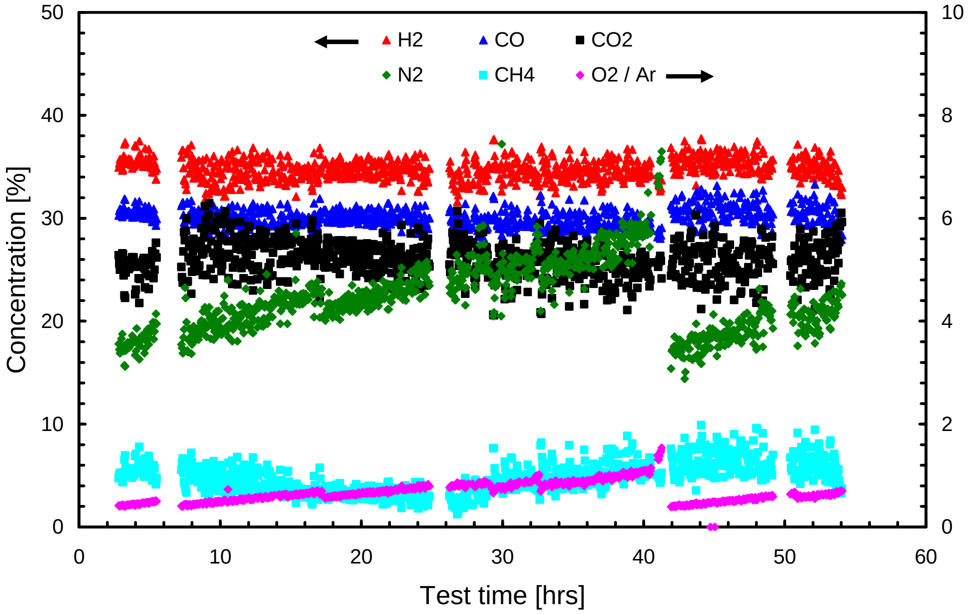

Figure 4.10*Composition of dry syngas from industrial waste downstream of tar cracker.*

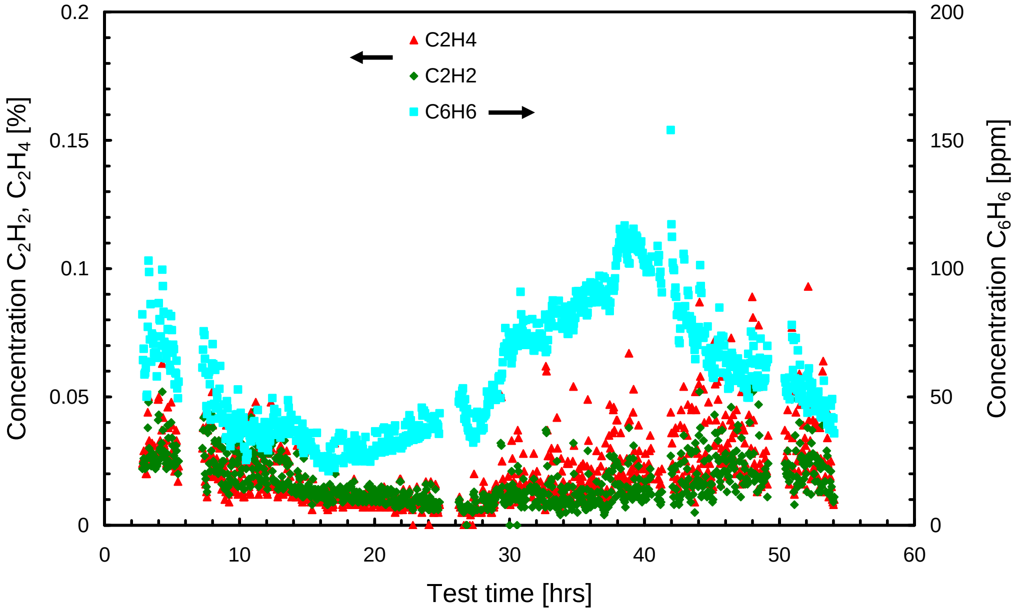

Figure 4.11*Concentration of C_2H_2, C_2H_4 and C_6H_6 in dry syngas from industrial waste downstream of tar cracker.*

SOFC performance

The SOFC stack was tested with constant 10 A load, except for interruptions due to problems or regular measurements of the impedance or I-V curves. The average stack voltage was 21.5 V, which means the stack delivered 215 W power. Figure 4.12 shows the measured stack voltage and voltages delivered by the 10 sets of three cells. Set 4 performs relatively poorly.

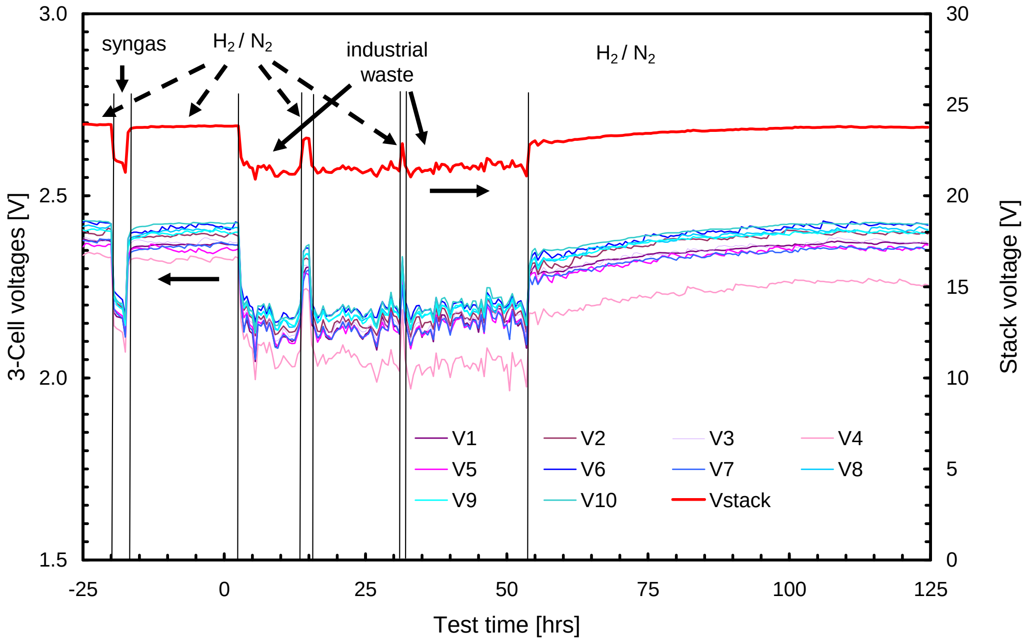

Figure 4.12*SOFC performance during test with syngas from industrial waste.*

Prior to the test, the stack delivered 219 W on synthetic syngas (water content 28%, dry gas composition 43% N_2, 23% H_2, 17% CO and 17% CO_2) and 238 W on H_2/N_2. The performance on syngas is slightly better than after the first test, which is remarkable given the N_2 dilution. The performance on H_2/N_2 is slightly worse than before. The fuel utilization calculated from the off-gas analysis and the C balance is 51.1%. A slightly higher value of 52.2% is calculated from the current drawn and the fuel input.

After the syngas test, the stack was fed with a gas mixture of 40% H_2 and 60% N_2. The output rose quickly to 230 W, followed by a gradual increase to 235 W after 24 hours and 238 W after 72 hours. The slow "recovery" is probably due to a change in stack temperature. The performance of set 4 remained poor in comparison with the other sets, a clear sign of degradation. The lower performance may already be visible before the test, but the difference grows over time. The degradation may have been caused by the presence of small quantities of C_2H_4 and C_6H_6 in the syngas, but may also be a sign that the stack is approaching its end of life after more than 3000 hours of operation.

Table 4.11 shows the average composition of syngas upstream and downstream the SOFC stack. The Ar value between brackets was derived by the method explained in Section 4.1 to make the Ar balance confirm to the C balance. Here, no correction is needed to the N_2 concentration to make the N balance agree with the C balance. As explained in Section 4.1, that points to air leaking into the syngas and not into the gas analysis system.

Table 4.11 *Average composition of dry syngas from industrial waste upstream and downstream of SOFC stack. Water content of syngas 36% upstream and 48% downstream of SOFC.*

		SOFC in	SOFC out
H_2	%	34.1	24.8
CO	%	27.3	14.2
CO_2	%	23.2	50.2
CH_4	%	0.90	0.01
N_2	%	5.7	7.5
O_2/Ar (Ar)	%	5.5 (4.2)	5.3

Based on the Ar content in the off-gas and the Ar flow added to the input gas, the dry off-gas flow would be 0.26 Nm^3/hr. From the C balance the dry gas input flow is calculated to be 0.326 Nm^3/hr. From the H balance, the off-gas water content is calculated to be 48%.

The dry syngas flow and composition correspond to a thermal input to the SOFC of 736 W (HHV) or 673 W (LHV). The dry off-gas flow and composition correspond to 357 W (HHV) or 322 W (LHV). The fuel utilization by the SOFC is 52%. The SOFC electrical efficiency is 29.2% (HHV) or 31.9% (LHV). The remaining 22.3% (HHV) or 20.2% (LHV) are converted into heat.

Discussion

As mentioned in the discussion at the end of Section 4.1, the calculated syngas flow may be too low and the electrical efficiency too high.

The low fuel utilization during the third test may also have been caused by the presence of air in the syngas, as explained at the end of Section 4.1. If the difference in N_2 concentration between the first two and third tests are due to air, the O_2 concentration would have been 0.5%. That would suffice to burn 2% of the syngas and bring the fuel utilization by the SOFC stack at 54% instead of 52%.

The difference in fuel utilization may just be an artefact of the analysis. During the first two tests, the off-gas composition was determined with a µGC. During the last test, a GC was used. At the end of the first test, fuel utilization with synthetic syngas was calculated to be 56.8%. Before the third test, fuel utilization with synthetic syngas was only 51.1%. In both cases the fuel utilization should have been 52.2% at the current drawn and fuel fed. As the results for syngas from beech wood and industrial waste differ by less than 1% from those of synthetic syngas, it may be concluded that the fuel utilization in all three endurance tests was probably close to 53%.

5. System analysis

The experimental results provide data which can be used to design a demonstration plant as the next step in the development. In order to bridge the gap between experiments at kW scale to an installation at MW scale, computer simulations have been performed with Aspen software. The simulations allow a system analysis of various technical options.

The simulations have been performed in a number of steps. First, a model of the experimental arrangement has been made. Results have been compared with those from experiments to validate the model. Next, the model has been used to determine the effect of various options and additional components to the system. Finally, the model has been used for the design of the 25 MW installation for which an economic analysis has been performed.

5.1 Basic model

The model operates essentially as a "black box" for the conversion of biomass or waste into electrical power and heat. Various other streams can go in or out of the system, as represented in Figure 5.1. The system operates at atmospheric pressure. Heat loss is neglected[3].

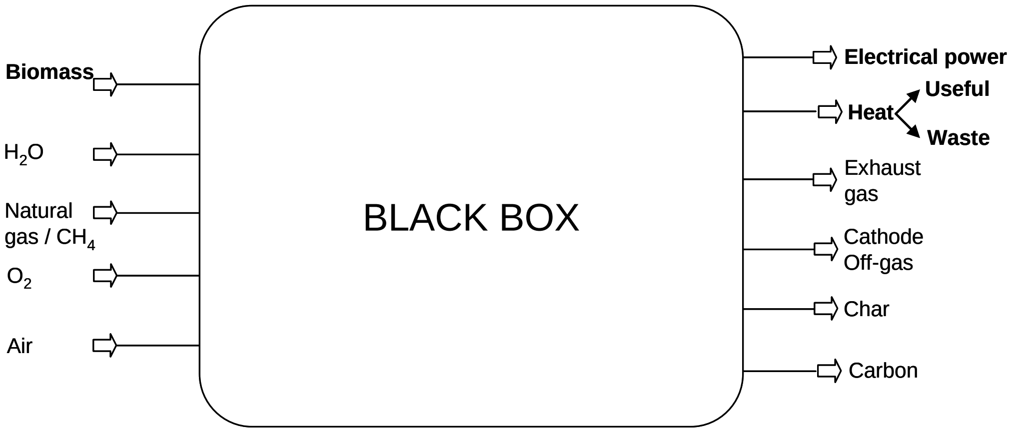

Figure 5.1 *Simplified scheme of a system for the conversion of biomass or waste into electrical power and heat by two-step gasification and fuel cells.*

5.1.1 Basic model building blocks

The black box shown in Figure 5.1 contains 6 blocks or conversion steps as indicated in Figure 5.2. Heat exchangers in between steps are not shown, but included in the model. The effect of gas cleaning on the heat balance is neglected. Other standard conditions used for the simulation of each step are given below.

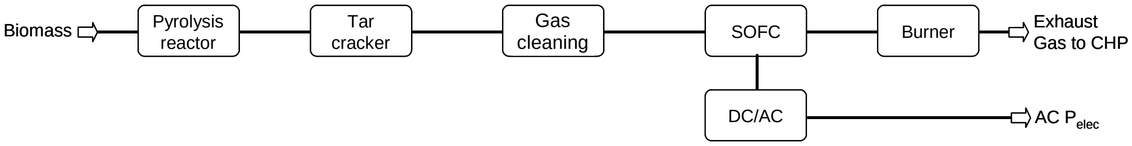

Figure 5.2 *Conversion steps within black box model of two-stage gasification and SOFC power production.*

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[3] Heat loss of the experimental installations is large and compensated by combustion of CH_4 or electrical trace heating. Heat loss is neglected in the simulations to allow extrapolation to a large installation in which heat loss will be limited.

Pyrolysis reactor
The pyrolysis reactor operates at 550°C. Fuel enters the pyrolysis reactor at 50°C. The feed rate is set at 3 kg/hr beech with 8% moisture for comparison with the experiments. The fuel composition is taken from experimental data. Steam is added at 200°C and a rate of 1 kg/hr. The steam is produced from 25°C water.

Pyrolysis gas and char leave the reactor at 550°C. The char composition is taken from experimental data. The char production is set at 15% of the fuel input. Pyrolysis gas consists of dry gas (H_2, CO, CO_2, CH_4 plus minor amounts of N_2, H_2S and HCl), water and tar. The tar composition is taken as 48% C, 8% H and 44% O by weight.

Tar cracker
The tar cracker converts pyrolysis gas into syngas (H_2, H_2O, CO and CO_2). Syngas leaves the reactor at 1125°C. The heat requirement is met by stoichiometric combustion of CH_4 with O_2. Both gases are provided at ambient temperature and pressure. Soot production is neglected.

In the experimental installation only part of the syngas is cleaned for use in the SOFC stack. In the simulation, the SOFC system is matched to the total syngas flow.

Gas cleaning (SACHA)
The gas cleaning operates at 300°C. The gas cleaning removes soot, H_2S and HCl, but the effect on the heat balance is neglected. Gas cleaning is included in the model to allow calculation of the amount of absorbents needed, which is relevant for the economical analysis.

SOFC
In the model, the size of the SOFC stack is chosen to match the available syngas flow. The SOFC stack operates at 750°C inlet temperature and 850°C outlet temperature for both syngas and air. A large excess of air is used to absorb the heat evolved in the conversion of syngas into electricity. The SOFC stack produces DC current at 0.7 V per cell at 60% fuel utilization. These values correspond to an electrical efficiency of 33.5% (LHV) for pure H_2 and 28.6% for pure CO.

DC/AC
The DC/AC converter produces 230 V AC current from the SOFC DC current at 97% conversion efficiency.

Burner
The afterburner produces heat by combustion of SOFC off-gas (depleted syngas) from the SOFC with ambient air at 20% air excess.

Auxiliaries
A pressure drop of 0.1 bar is assumed to calculate the power requirement of the anode gas blower and of the air blowers for the SOFC and the afterburner. The isentropic efficiency of the blowers is set at 65% and the mechanical efficiency at 95%. Energy for O_2 production is not taken into account.

5.1.2 Heat integration

At various points in the system heat has to be provided or removed. The main heat sinks and sources are given in Table 5.1 and Table 5.2. Some of the sinks and sources are unlikely to be realized in practice. Preheating of pyrolysis gas will lead to coke formation and blocking of heat exchangers. Cooling of syngas upstream of the gas cleaning will require expensive materials if fuels are used with high Cl or S content. Drying of wet biomass is not included, but can be an attractive option to use heat from the process and improve the system efficiency. Char cooling and further use of char are not relevant at this point, but will be discussed at a later stage.

Table 5.1 *Heat sinks in two-stage gasification and SOFC system with entrance and exit temperatures.*

Heat sink	*T in [°C]*	*T out [°C]*
Water => steam	25	200
Biomass => pyrolysis gas	50	550
Pyrolysis gas preheating	550	> 550
Pyrolysis gas => syngas	> 550	1125
Clean syngas => SOFC	300	750
Air => SOFC	25	750

Table 5.2 *Heat sources in two-stage gasification and SOFC system with entrance and exit temperatures.*

Heat source	*T in [°C]*	*T out [°C]*
Syngas cooling (before cleaning)	1125	300
SOFC off-gas cooling	850	> 25 [1]
SOFC O_2 depleted air cooling	850	> 25 [1]
CH_4 and/or pyrolysis gas combustion	25 / 550	1125 [2]
SOFC off-gas combustion	> 25 [1]	… [3]

[1] Temperatures depend on amount of heat withdrawn in cooling.

[2] Temperature after mixing with pyrolysis gas and cracking of hydrocarbons.

[3] Temperature depends on gas utilization and combustion conditions.

A heat exchanger model allows analysis of various options to cover heat demand by heat supply within the system and serves to calculate the amount of heat available for export out of the system. The Aspen MheatX module has been used to perform a pinch analysis of all heat demand and supply streams. The data have also been used in the Odessy PRO pinch analysis program to construct a heat exchanger network that is optimal from an economical point of view. The results have been used to design the system shown in Figure 5.3.

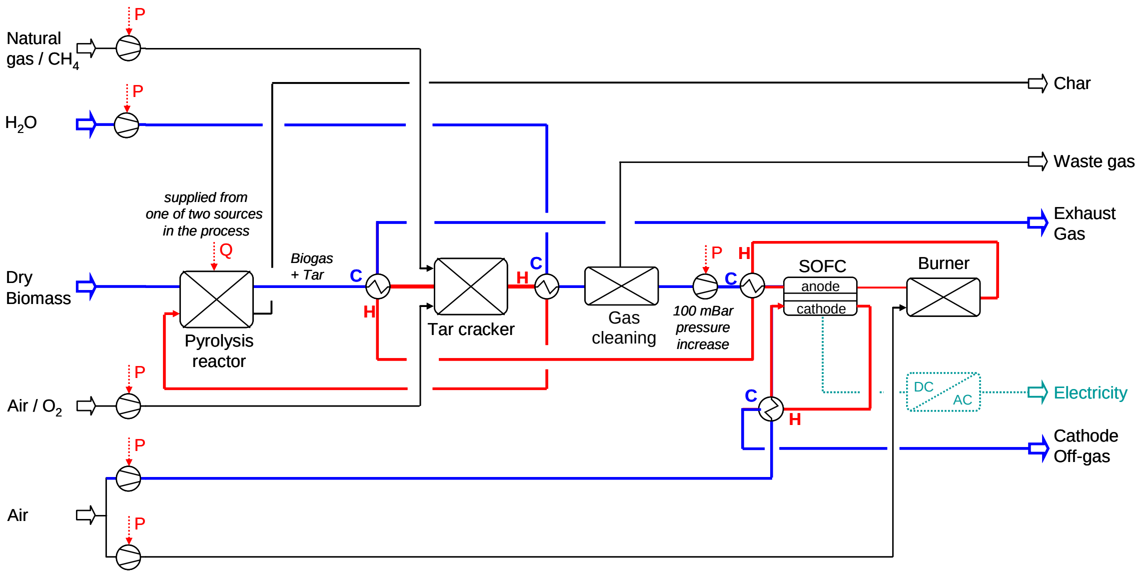

Figure 5.3 *Heat exchange network design for two-stage gasification and SOFC stack, with hot (H) and cold (C) streams, heat (Q) and electrical power (P) input.*

Syngas from the tar cracker is used to produce steam for the pyrolysis process. The syngas contains sufficient heat to produce steam for other purposes too. It can also be used to heat the pyrolysis reactor, i.e. it is one of the two sources mentioned in Figure 5.3.

Fresh air for the cathode of the SOFC stack is heated to 750°C by O_2 depleted air which leaves the SOFC stack at 850°C. The cathode off-gas cools to about 150°C. It can still provide low-temperature heat, e.g. for biomass drying.

Off-gas leaving the anode of the SOFC stack at 850°C can be burned with air to produce high temperature flue gas. The exhaust gas is used to heat clean syngas to the SOFC entrance temperature of 750°C. In the network shown, it is also used to heat pyrolysis gas entering the tar cracker. The exhaust gas remains hot enough to provide high quality heat for other purposes, e.g. steam production.

5.1.3 Basic model results

Base case, methane combustion in tar cracker

According to the model, the input of 3 kg/hr beech wood and 1 kg/hr steam yields 0.45 kg/hr char and 3.55 kg/hr pyrolysis gas. The energy content of the char corresponds to 26.5% of that of the beech wood. That is significantly less than the 42% found in the beech wood test, but in line with the experimental results of 19% and 24% for Rofire and industrial waste.

The pyrolysis gas contains 1.35 kg/hr water, 0.71 kg/hr tar and 1.49 kg/hr dry gas. The water content of the pyrolysis gas is 59%, to be compared with an experimental value of 77% (see Section 4.1). The calculated tar content is 610 g/Nm^3, to be compared with a measured value of 730 g/Nm^3. Apparently, the model predicts reactions between water and tar that yield more dry gas than observed in practice.

Heating of the pyrolysis gas in order to crack tar and produce syngas at 1125°C requires the combustion of 0.36 Nm^3 CH_4 with 0.72 Nm^3 O_2. The large difference with the experiment shows how important heat loss can be for a small installation. The calculated syngas contains 30% moisture and has a dry gas composition of 50% H_2, 35.7% CO and 14.3% CO_2. The actual syngas contained 43% moisture and had a dry gas composition of about 38% H_2, 31% CO and 30% CO_2 (from Table 4.5, corrected for inert gases). The larger moisture and CO_2 content are due to CH_4 burned to compensate heat loss.

According to the model, the net heating value of the syngas is 85% of the net heating value of the wood input. It is only 67.5% of the total heat input from wood plus methane. At 3 kg/hr wood input, the net heating value of the syngas is equivalent to 11.8 kW. At 60% fuel utilization in the SOFC, the SOFC off-gas flow is equivalent to 4.7 kW. The off-gas contains 50% moisture and has a dry gas composition of 52% CO_2, 30% H_2 and 18% CO. Again, the experimental values for H_2O and CO_2 are a little higher.

The SOFC produces 3.67 kW$_e$. Taking into account conversion loss in the DC/AC conversion (0.11 kW$_e$) and power use by compressors (0.48 kW$_e$), the calculated net electricity production becomes 3.08 kW$_e$. The net efficiency is 22.2% with respect to the beech wood fuel input and 17.6% with respect to the total fuel input of wood plus methane.

Base case, syngas combustion in tar cracker

A commercial large scale installation would preferably not use methane in the tar cracker. Instead, part of the pyrolysis gas or syngas would be burned. Because of the combustion of part of the H_2 and CO, the water content of the syngas would increase to 34.5% and the dry gas composition would become 45.9% H_2, 35.3% CO and 18.8% CO_2. The net heating value of the syngas corresponds to 8.04 kW, i.e. 58% of the net heating value of the wood fuel.

The SOFC produces 2.70 kW$_e$. Loss in the DC/AC converter (0.08 kW$_e$) and power use by compressors (0.32 kW$_e$) reduces the net power output to 2.30 kW$_e$ and brings the net electrical efficiency at 16.6%. The 1% efficiency loss with respect to the base case with methane use could be recovered if pyrolysis gas were preheated to 900°C by heat exchange with hot syngas leaving the tar cracker. The main obstacle would be soot formation in the heat exchanger.

Analysis

The basic model results in a system efficiency from biomass to electricity of only 17%. The remaining 83% are released as heat or converted into char and SOFC off-gas with considerable heating value. An analysis of the efficiency of separate steps is a useful basis for optimization.

- The first gasification stage has an efficiency of 73.5% from wood to pyrolysis gas. The remaining 26.5% are bound in char.
- The second gasification stage has an efficiency of 79% from pyrolysis gas to syngas. The remaining 21% are released as heat. The syngas heating value is 58% of the wood energy content.
- The SOFC utilizes only 60% of the syngas. The remaining 40%, equivalent to 23% of the wood energy content, is still available as depleted syngas.
- The syngas used by the SOFC is converted into electricity with 56% efficiency. The remainder is released as heat.
- The DC/AC converter and gas compressor require 15% of the SOFC electricity production.

According to this analysis, char and depleted syngas together contain nearly 50% of the wood energy content. About 30% is converted into heat in high-temperature processes. The gross electrical efficiency of 20% is reduced to a net efficiency of 17% by the system power use.

5.2 System improvement options

The net electrical efficiency for the base cases presented in the previous section is too low to make the proposed system a viable option. There are several options to increase the efficiency. The options discussed in this section are: more efficient use of syngas, use of excess heat for electricity production, use of char, and reduction of system power use.

5.2.1 Fuel utilization

80% fuel utilization

The base cases assume 60% fuel utilization. That is considered to be a safe operating condition of this specific SOFC stack design, but results in too low an electrical efficiency. Experiments performed as part of the previous project show that fuel utilization can be increased to 80% of even more for other SOFC stacks [1]. The test reported at the end of Section 4.2 proves that a fuel utilization of about 75% is possible with syngas produced from Rofire in a commercial SOFC stack, albeit at a 10% lower SOFC stack performance. The low performance is due to the design of the stack, which results in an uneven flow distribution and internal by-pass of fuel.

The lower stack performance is caused by the fact that part of the SOFC stack or part of each SOFC cell operates with syngas from which most of the H_2 and CO has been converted. That results in increased polarization loss and lower output voltage. That way, part of the efficiency gain from the increased fuel utilization is lost again.

A higher output voltage from depleted syngas can be obtained at a lower current density. Several SOFC stacks connected in series to the syngas supply could operate at the same voltage but at progressively lower current densities. The total power output increases proportionally to the fuel utilization. Auxiliary power use would grow too, but a little less[4]. If the fuel utilization can be increased from 60% to 80%, the net electrical efficiency increases from 17.6% to 23.7% for the base case with methane use and from 16.6% to 22.4% for the base case without methane use in the tar cracker.

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[4] The SOFC air compressor is responsible for 80% of the system power use. The air flow it provides has to increase with power output to remove heat generated as a by-product. Hence, its power use is in proportiion to the power output. The syngas compressor, the second biggest power user, provides a syngas flow independent of the fuel utilization. Its power demand remains constant, except when higher pressure is required to supply SOFC stacks connected in series. The air compressor of the afterburner, which takes about 5% of the system power use, requires less power with increasing fuel utilization, as less air is required to burn leaner off-gas.

Operating a stack at reduced current density means a larger surface, i.e. a larger and more expensive stack, is required. The additional investment costs have to be balanced with the proceeds from the net gain in power output.

Low-temperature gas cleaning
High-temperature gas cleaning was mentioned in the proposal for the present project as one of the options to improve system efficiency and reduce costs. It reduces the amount of syngas cooling and heating required upstream and downstream of the gas cleaning. It would also reduce the heat loss associated with each cooling or heating step.

High-temperature gas cleaning has the known disadvantage that it is more difficult to remove contaminants by absorbents to the low levels required for SOFC operation. Still, the experiments reported in Chapter 4 showed gas cleaning to be sufficient, at least to the point that no signs were observed of degradation in SOFC performance over 250 hours of operation on syngas cleaned by commercial absorbents at 200°C.

The above discussion on fuel utilization makes evident, that high-temperature gas cleaning has another disadvantage: steam added to suppress soot formation or produced in the two-stage gasification process remains in the syngas and hinders diffusion of H_2 and CO to the anode surfaces of the SOFC stack. Although some water is required to allow CO conversion to H_2 by the shift reaction and to prevent soot formation in the SOFC, 30% to 35% (see Sub-section 5.1.3) is far too much.

Reduction of the water content to about 5% requires cooling to 35°C, a temperature level which can be reached by cooling towers with moderate power requirements. The efficiency burden is less than 0.5% with respect to the wood fuel input. That figure should be compared to the 6% gain if the fuel utilization could be increased from 60% to 80% because of the lower water content.

Removal of most of the water reduces the amount of heat required to raise the temperature of the cleaned syngas to the SOFC entrance temperature of 750°C. In fact, heating dry syngas from 35°C to 750°C requires no more heat than heating the equivalent amount of wet syngas from 300°C to 750°C. Syngas drying has the additional advantage that combustion in the afterburner yields a smaller volume of hotter exhaust gas. It means, heat can be recovered more easily and effectively. The use of heat for additional power production is discussed below.

5.2.2 Steam bottoming cycle

As mentioned in Sub-section 5.1.2, there are several heat streams that contain heat which could be used to increase the system electrical efficiency. Some efficiency gain can be realized by increasing the temperature of steam used in the pyrolysis unit. Pre-heating pyrolysis gas is another option, discussed already at the end of Sub-section 5.1.3.

The only substantial contribution can be expected from direct use of heat for power production. The most promising option is the integration of an SOFC stack and a gas turbine. It involves operating an SOFC stack at above atmospheric pressure, combustion of the SOFC off-gas with O_2 depleted air from the SOFC stack and expansion of the exhaust gas over a gas turbine. Very high efficiency is expected from such a system, up to 70% for MW systems using natural gas. Siemens Westinghouse has demonstrated the feasibility of that approach at 220 and 300 kW_e scale, but acknowledges that these systems are quite complex and costly. Because of the feasibility and potential further development is warranted, but hybrid systems with fuel cells at atmospheric pressure will be examined first [3].

At present, the most realistic option is the use of excess heat to produce steam which is expanded in a steam turbine driving a generator. For small installations, the use of an organic Rankine cycle can be considered instead. Here, a condensing steam bottoming cycle is considered (see Figure 5.4). Water under pressure is heated to 105°C by cathode off-gas (O_2 depleted air). The off-gas temperature is 142°C because it has been used to heat ambient air to the SOFC entrance temperature of 750°C (see Figure 5.3). Afterburner off-gas (the combustion product of syngas after utilization in the SOFC) is used first to heat syngas and pre-heat pyrolysis gas (see Figure 5.3), and then to produce overheated steam. The off-gas temperature depends on the SOFC fuel utilization. Steam is expanded to 0.04 bar over a turbine with 90% isentropic efficiency. The steam turbine drives a generator with 97% efficiency. Power use for cooling of the condenser is set at 5% of the heat to be removed.

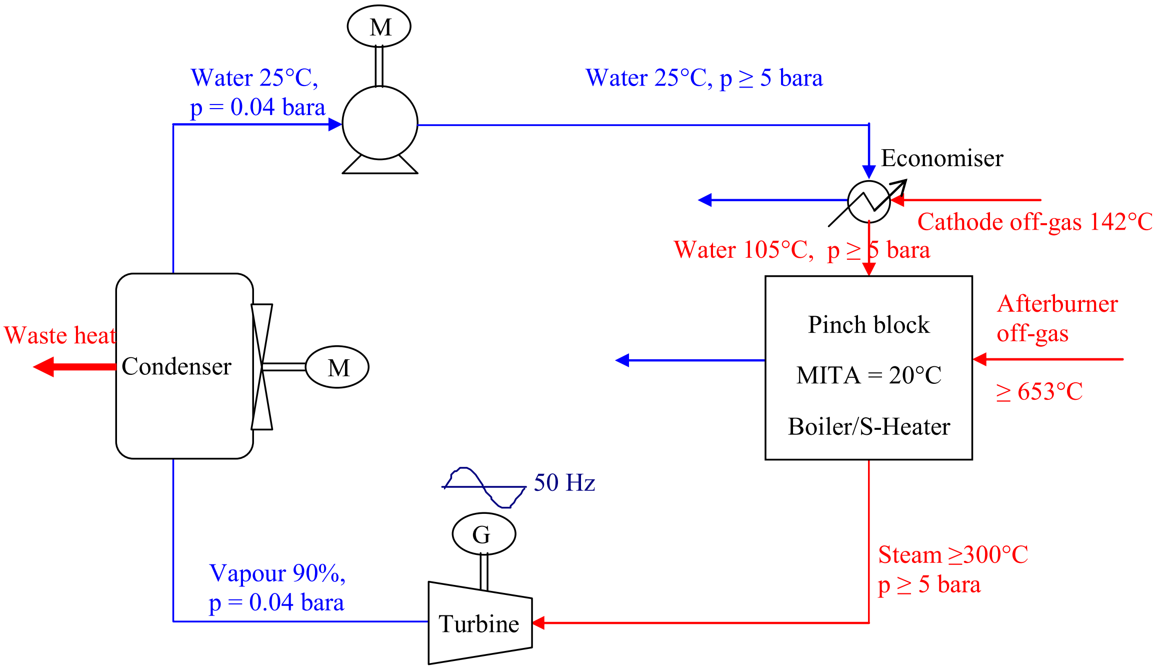

Figure 5.4 *Steam bottoming cycle using heat from SOFC stack and afterburner.*

The steam pressure and temperature have been varied to find the optimum conditions for 60% and 80% fuel utilization by the SOFC stack. Combinations of pressure and temperature are used which result in a 90% vapour fraction after expansion to 0.04 bar. Table 5.3 shows results for systems using methane to heat the tar cracker. Table 5.4 shows results for systems in which no methane is used but the pyrolysis gas preheated to 900°C. Clearly, the steam cycle is most useful at low fuel utilization by the SOFC stack, when the off-gas contains most energy. Similarly, the steam cycle adds relatively little when off-gas has been used to preheat the pyrolysis gas. Calculations for other conditions show that system efficiency changes little when the pyrolysis gas temperature is changed. At higher temperature the SOFC gets richer syngas and produces more power, but less energy is left for the steam cycle. At lower temperature or without preheat, the reverse is true.

Table 5.3 *Efficiency of two-stage gasification with SOFC and steam bottoming cycle (SBC) for systems with 13.88 kW wood and 3.6 kW methane fuel input.*

Fuel utilization		*60%*	*60%*	*80%*	*80%*
Steam temperature	°C	500	600	500	600
Steam pressure	bar	27	56	27	56
SBC gross power	W_e	1978	2180	1158	1277
SBC power use	W_e	203	195	119	113
SBC net power		1775	1985	1039	1164
SBC efficiency	%	10.2	11.4	6.0	6.7
SOFC net power	W_e	3080	3080	4140	4140
System net efficiency	%	27.8	29.0	29.6	30.4

Table 5.4 *Efficiency of two-stage gasification with SOFC and steam bottoming cycle (SBC) for systems with 13.88 kW wood. Pyrolysis gas preheated to 900°C.*

Fuel utilization		*60%*	*60%*	*80%*	*80%*
Steam temperature	°C	500	600	500	600
Steam pressure	bar	27	56	27	56
SBC gross power	W_e	1182	1304	501	532
SBC power use	W_e	122	118	51	47
SBC net power		1060	1186	450	485
SBC efficiency	%	7.6	8.5	3.2	3.5
SOFC net power	W_e	2470	2470	3330	3330
System net efficiency	%	25.4	26.3	27.2	27.5

5.2.3 Char use

The combination of an SOFC stack and a steam bottoming cycle can bring the net system efficiency for power production near 30%. If no methane is used, the efficiency can be around 27%. That figure becomes more impressive if it is realized that about one quarter of the energy content of the wood fuel is left in char and not utilized. The fuel fraction that is used is converted into power with 37% efficiency.

If a substantially higher efficiency is required, char has to be used for energy production. Char can be burned to produce steam for the steam bottoming cycle, or char can be converted by steam gasification to syngas for use in the SOFC stack. In both cases, additional gas cleaning will be required as trace elements and contaminants in the fuel fed to the pyrolysis unit are concentrated in char.

The combustion of char for steam production could increase the system efficiency by 7% if the steam cycle would operate at 30% efficiency. Steam gasification of char produces syngas which can be utilized with 37% efficiency as shown above. Hence, the system efficiency could increase to 37%. In fact, it can be even higher as steam gasification of char is an endothermic process which is able to convert heat into chemical energy. In the next chapter on the economics of large systems, char gasification is included in the analysis.

5.2.4 Auxiliary power consumption

The results presented above show that a substantial part of the power produced is consumed by ancillary equipment. In case of the steam bottoming cycle, there is little room for reduction of auxiliary power consumption as the technology is mature. For the SOFC stack, a relatively simple configuration has been considered, which still leaves room for improvement.

As mentioned above, about 80% of the power consumption by ancillary equipment is taken by the SOFC air compressor. It is equivalent to 2% of the fuel input to the pyrolysis unit for the base cases. It can reach 3% if char gasification and high fuel utilization are realized.

Only a fraction of the air provided is needed for the chemical reactions. Much more is needed for cooling of the SOFC stack. The amount depends on the allowable difference between entrance and exit temperatures. A value of 100°C is used in the present calculations. That value is a compromise between SOFC stack performance, stack lifetime and power consumption.

The amount of air needed can be reduced if SOFC stacks are connected in series, with cooling in between stacks. Series connection of stacks would increase the pressure drop to be overcome by the compressor. The total pressure drop would grow less than linear in the number of series-connected stacks as there would be less heat exchanger surface involved. The use of O_2 depleted air would only marginally affect the SOFC stack performance if only two or three stacks were connected in series.

Series connection of SOFC stacks could reduce auxiliary power consumption and increase the system efficiency considerably. Series connection of SOFC stacks has other advantages that may be equally important. In the system shown in Figure 5.3, series connection reduces the required air compressor capacity and the size of heat exchangers between fresh air and exhaust air. Even though additional heat exchangers are needed for cooling in between stacks, the total cost will go down. The effect on system economy will be discussed in the next chapter.

6. Demonstration plant design

The results of the optimization and endurance tests reported in Chapters 3 and 4 show that two-stage gasification of biomass and waste can produce syngas of sufficient quality for SOFC operation. The commercial SOFC stack used showed no degradation during tests in which three different fuels were used for 250 hours in total. The stack showed only minor degradation over a period of more than 3000 hours in which the stack operated with a H_2/N_2 mixture [4].

The SOFC stack used was obtained from Staxera GmbH. Commercial production of SOFC stacks by Staxera started only recently. The stack design and production process are subject to continuous development and improvement. Current stacks show less than 0.4% performance degradation per 1000 hours on full load. More recent stacks can deliver 50% more power. Staxera has now also developed an integrated SOFC module ready for connection to gas supply, sensors and current leads. It delivers 1 kW_e at 75% fuel utilization, shows less than 1% degradation per 1000 hours and is guaranteed to withstand at least 20 thermal cycles [5].

At present, both the size and price of SOFC stacks and systems make them unsuited for commercial application in systems which are the subject of the present project. Staxera intends first to develop, manufacture and sell SOFC stacks of 1 to 5 kW_e size for household and mobile applications. Topsøe Fuel Cell and Wärtsilä are also developing SOFC systems for distributed power generation. They have demonstrated more than 4000 hours of operation of a 5 kW_e SOFC system. A 20 kW_e unit was taken into operation in 2007, a 50 kW_e system is planned for 2008 and a conceptual study was made of a 250 kW_e system [6, 7]. According to the latter study, SOFC units should reach a turnkey price level of 1589 €/kW between 2010 and 2015.

While Staxera and Topsøe Fuel Cell use planar fuel cells, Siemens Westinghouse focuses on tubular fuel cell technology. A 100 kW_e system has been demonstrated and is a 250 kW_e system is in development. However, Siemens also develops planar fuel cells in a project funded by the Solid State Energy Conversion Alliance (SECA). Targets of the SECA program are to reach a price level of 400 $/kW for a complete system and 0.1% degradation per 1000 hours.

For the design and analysis of a demonstration plant we have assumed SOFC systems will become available in MW size at a price level of 1600 €/kW_e. ECN has performed a study of a 25 MW size plant based on the technology used in the endurance tests described in Chapter 4 and analysed in Chapter 5. Techno Invent has performed a study of a 2.5 MW size plant using a different technology called Green MoDem. The latter approach involves thermal conversion of solid fuel into oil and syngas. The syngas is used in an SOFC system. Oil and SOFC off-gas are used in a dual fuel engine.

6.1 25 MW system using two stage gasification

6.1.1 System description

A 25 MW demonstration plant has been designed for the use of Rofire fuel as in the second endurance test (see Section 4.2). It requires 3850 kg/hr Rofire fuel with 1.6% moisture, to which 3730 kg/hr steam is added for a 1:1 ratio of steam and water to dry fuel. Starting point for the design was the two-stage gasification system with SOFC and steam bottoming cycle described in Sub-section 5.2.2, extended with gasification of char as described in Sub-section 5.2.3. The process flow diagram is shown in Figure 6.1. Further improvements as described in Sub-section 5.2.4 have been included at a later stage and will be discussed at the end of Sub-section 6.1.2.

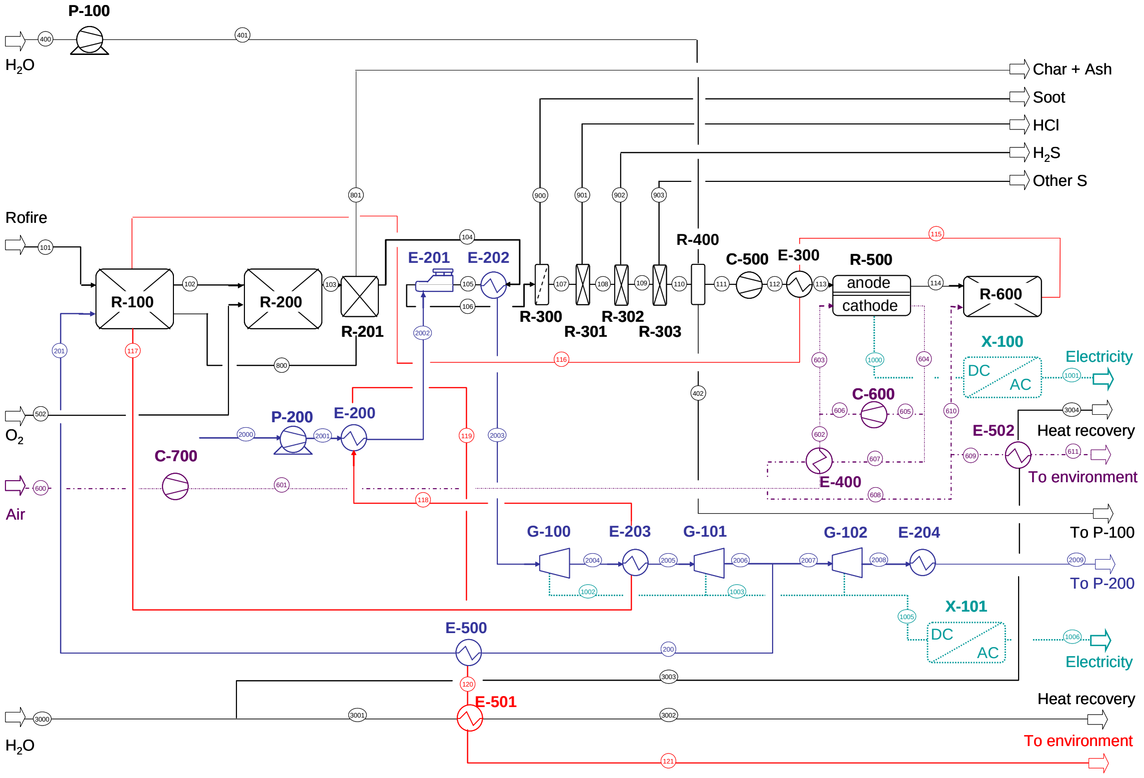

Figure 6.1 *Process flow diagram for a 25 MW demonstration plant.*

Rofire fuel and steam are fed to the pyrolysis reactor (R-100), which is heated by afterburner (R-600) exhaust gas. The tar cracker (R-200) is heated by combustion of part of the pyrolysis gas with O_2. Char from the pyrolysis reactor and syngas from the tar cracker are fed to the char gasifier (R-201). In there, CO_2 and H_2O from the syngas react with char to form additional CO and H_2. Due to the endothermal nature of the char gasification reactions, the syngas cools from 1125°C to 900°C. The syngas is cooled further in the steam superheater (E-202) and steam generator (E-201). Soot, Cl and S are removed by a filter (R-300) and reactors filled with sorbent materials (R-301 to R-303). The gas is dried to 5% moisture by cooling (R-400), compressed (C-500), heated (E-300) and fed to the anode of the SOFC stack (R-500).

The SOFC stack operates at 80% fuel utilization and 0.7 V/cell. Conversion from DC to AC power (X-100) is performed at 97% efficiency. Anode off-gas is burned in the afterburner (R-600). Afterburner exhaust gas is used to heat syngas (E-300), the pyrolysis reactor (R-100), reheat steam (E-203), preheat water (E-200) and reheat low pressure steam (E-500) for use in the pyrolysis reactor. High-quality heat can still be recovered (E-501). For the economic evaluation heat export is assumed to a system in which water is heated from 70°C to 90°C

Air is compressed (C-700), preheated (E-400), mixed with part of O_2 depleted air, and fed to the cathode of the SOFC cathode (R-500). Part of the O_2 depleted air from the SOFC cathode is recompressed (C-600) and mixed with preheated fresh air. The remainder is used to preheat fresh air (E-400). Some of it is then used in the afterburner (R-600), most of it is vented to the environment, after recovery of heat (E-502). Again, heat of high quality can be obtained, but for the economic evaluation export to a 70°C/90°C hot water system is assumed.

The steam cycle contains three expansion steps (G-100, G-101 and G-102) with reheat (E-203) after the first expansion step and steam extraction for use in the pyrolysis unit after the second expansion step. After the last expansion step, steam is condensed (E-204) and the water recycled. Water lost due to steam extraction can be made up by fresh water and by water removed from the syngas in the drier (R-400). The steam turbines have an isentropic efficiency of 90%, the AC generator (X-101) an efficiency of 97%.

6.1.2 System performance

Process mass flows

The fuel input to the 25 MW system is 3850 kg/hour Rofire pellets. Other inputs are 3730 kg/hr steam and 1660 kg/hr O_2 for the tar cracker, and 75 tons/hr air to the SOFC cathode. The system produces 300 kg/hr ash which still contains 15% carbon. It is assumed that 1/3rd of S and Cl in the Rofire are bound in the ash. The remainder is converted into H_2S and HCl and removed from the syngas by the gas cleaning. That yields waste streams containing nearly 2 kg/hour S and 50 kg/hr Cl. The syngas dryer removes 2380 kg/hr water which can be re-used for steam production. The SOFC anode receives 6500 kg/hr syngas with a net heating value of 12.2 MJ/kg which is equivalent to 22 MW. The anode off-gas flow is 10500 kg/hr and equivalent to 4.4 MW. It is burned in the afterburner with 6350 kg/hr O_2 depleted air from the SOFC cathode. Exhaust gas from the afterburner and the remaining O_2 depleted air from the SOFC cathode are vented to the atmosphere.

Process energy flows

The 25 MW fuel input is converted to 22 MW syngas. At 80% fuel utilization the SOFC stack produces 9.3 MW_e DC power. After conversion to AC power and subtraction of auxiliary power consumption, a net power output of 7.5 MW_e remains. Total heat input to the steam cycle is 5.1 MW. The steam bottoming cycle produces a net power output of 1.3 MW_e. That brings the total system output at 8.8 MW_e and the net electrical efficiency at 35.2%[5].

Heat recovered from the afterburner exhaust gas (E-501) and from the O_2 depleted air (E-502) is transferred to a system in which water is heated from 70°C to 90°C. The afterburner exhaust gas can deliver 3.5 MW and the O_2 depleted air 6.4 MW. The total of 9.9 MW corresponds to a thermal efficiency of 39.6%.

System improvement

As discussed in Sub-section 5.2.4 the system efficiency can be improved by reduction of the auxiliary power consumption. The main power consumer of the system shown in Figure 6.1 is the cathode air recycle compressor which requires nearly 1.3 MW_e. The air recycle is not needed if SOFC stacks are connected in series to the syngas and air supply[6]. However, the syngas and fresh air compressors would have to overcome a larger pressure drop and use more power. An increase by a factor 4 would require an extra 0.5 MW_e. The net saving would still be 0.8 MW_e. The reduced power consumption leads to a similar reduction of the amount of heat available in syngas and air for the steam cycle and for export. The net electrical efficiency will increase by 3% to 38.2% and the thermal efficiency will decrease by 3% to 36.6%.

The operating voltage determines the electrical and thermal efficiency. A higher value, e.g. 0.75 V instead of 0.7 V, would increase the electrical efficiency of the SOFC by 7%. It would also reduce the heat output and auxiliary power consumption for cooling. However, it would imply operation at lower current and power density, i.e. require a larger active area. That would result in higher SOFC investment costs, which will have to be matched by higher revenues from the additional electricity production. The value of 0.7 V/cell is considered the lowest safe operating voltage[7]. It is used in the economic analysis because it would minimize the SOFC investment costs[8].

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[5] As O_2 has been treated as a commodity, electricity consumption for O_2 production has not been taken into account.

[6] The SOFC stacks have to be connected in series to both the air and syngas supply to maintain the pressure balance between the anode and cathode sides.

[7] At lower voltage there is an increased risk of damage to the electrodes.

[8] For an analysis of the effect of the operating voltage on the system economy, the SOFC price would have to be specified per m^2 instead of per kW_e output.

6.1.3 System economy

The economy of the 25 MW system described above has been evaluated using the basic assumptions specified in Table 6.1. The total investment has been estimated from the bare equipment costs. Data are based on the most recent information from the Dutch Association of Cost Engineers (DACE) and information from www.matche.com. Data have been cross-checked with the ICARUS cost analysis program from Aspen Technology. However, the system contains several parts which are far from common and for which reliable data do not exist. For those parts cost estimates from other studies have been used.

The bare equipment costs of all parts except the SOFC system and steam turbine have been multiplied by Lang factors to account for the other direct and indirect investment costs (see Table 6.2). The costs of the SOFC system and steam turbine have been multiplied only by a factor 1.05 to account for connection to the electricity grid and the gas and steam supply. The latter assumption is in line with the approach followed for the calculation of the subsidy required to make power generation from renewable sources marginally profitable [8].

Table 6.1 *Basic assumptions for economic evaluation*

Economic parameter	*Value*	*Unit*
Power plant life	15	Year
Load factor	85	% (7500 hr/yr)
Heat utilization factor	50	% (3750 hr/yr)
Interest	10	%/year
Tax & insurance	2	%/year of plant investment including electric generators
Maintenance costs	4	%/year of plant investment excluding electric generators 15 €/MWhr net production for electric generators [1]
Number of operators	15	5 shifts of 3 persons each
Operator cost	60	k€/person.yr
Fuel (biomass) price	-0.823	€/GJ (@ -20 €/tonne)
Gas price	7	€/GJ (= 25 €/MWh)
Electricity (market) price	70	€/MWh
Adsorbent price	130	€/tonne
Oxygen price	70	€/tonne
Water price	1.31	€/m^3
Ashes disposal price	150	€/tonne

[1] The SOFC system requires comparatively little maintenance, but the stack will have to be replaced after 40,000 to 80,000 hours of operation.

It soon became clear that the cathode air recycle compressor (C-600) in the system shown in Figure 6.1 would be far too expensive because of the need to use high-temperature resistant Inconel material. The first alternative considered, a larger fresh air compressor (C-700) and heat exchanger (E-400) to supply the full cathode air flow at 750°C, was still very expensive and reduced the heat export capacity by 30%. Finally, the economic evaluation was performed for a system in which three SOFC stacks are connected in series, as discussed above. Table 6.3 summarizes cost data for that configuration.

In case of the gasification system, the price of the pyrolysis reactor (R-100) is the most difficult to estimate. Here, it has been obtained by comparison with a rotary kiln incinerator and results from previous studies [1, 9]. In case of the SOFC system, a price of 1600 €/kW$_e$ has been used for a complete system to be realized between 2010 and 2015 [7].

Application of the Lang factors as given in Table 6.2 brings the total investment costs for the plant excluding electric generators at 4.26*5.9 = 25.15 M€. The total investment for the plant including the electric generators amounts to 25.15+14.9*1.05 = 40.8 M€. At a net power output of 9.55 MW_e, the specific investment costs are 4275 €/kW_e. The capital required to cover costs during the construction and start-up phase adds another 20% to the investment. That brings the final total at 1.2*40.8 = 49 M€ and the specific investment costs at 5125 €/kW_e.

Table 6.2 *Lang factors for various direct and indirect costs.*

Cost parameter	*Factor*		*Multiplier*
Other direct investment costs			
Installation		0.47	Total bare cost of equipment
Instrumentation		0.18	Total bare cost of equipment
Piping		0.66	Total bare cost of equipment
Electrical equipment		0.11	Total bare cost of equipment
Building		0.18	Total bare cost of equipment
Service facility		0.15	Total bare cost of equipment
Total of direct investment costs	2.75		Total bare cost of equipment
Indirect investment costs			
Engineering and supervision		0.33	Total direct investment costs
Construction and supply		0.07	Total direct investment costs
Contractor's fee		0.05	Total direct investment costs
Contingency		0.10	Total direct investment costs
Total of indirect investment costs	0.55		Total direct investment costs
Total investment costs	4.26		Total bare cost of equipment

Table 6.3 *Bare equipment cost estimates.*

Identifier	*Equipment*	*Bare cost [k€]*	*Remark*
R-100, 200, 201	Gasification system	4,000	
R-300 to 303, R-400	Gas cleaning and drying	900	Soot, Cl and S removal
E-200 to E-204	HRSG	420	Heat recovery steam generator
E-500	Steam superheater	60	Pyrolysis steam
E-501	Flue gas heat recovery	200	Heat (hot water) delivery
E-502	Exhaust air heat recovery	300	Heat (hot water) delivery
P-100, P-200	Water pumps	20	
	Bare cost of equipment	**5,900**	**k€**
R-500, R-600, C-500, C-700, E-300, E-400, E-401, E-402, X-100	Complete SOFC system	14,400	9 MW_e net output
G-100 - 102, X-101	Steam turbine	500	Three stages, condensing
	Cost of electric generators	**14,900**	

Yearly revenues

If electricity is valued at 70 €/MWh, the yearly revenue from 7500 hrs/yr power production amounts to 5.01 M€. At a heat price of 7 €/GJ and 3750 hrs/yr utilization, revenue from heat supply amounts to 0.86 M€. The assumed negative fuel price of -20 €/ton adds another 0.58 M€. The total income from these three sources is 6.45 M€.

Yearly operational costs

Maintenance and insurance together amount to 2.9 M€/yr. Operation by a crew of 15 persons in 5 shifts leads to yearly costs of 0.9 M€. Yearly costs for O_2 amount to 0.87 M€. Ash disposal (0.34 M€), water supply and Cl and S absorbents (0.08 M€) bring the total yearly operational costs to 5.09 M€.

Profitability

The difference of 1.36 M€ between the yearly revenue and operational costs is insufficient to pay the 4.9 M€ rent on the total capital invested, let alone to repay the investment and make the plant profitable. That would require at least 5.08 M€ higher revenues or lower costs. The gap can be closed if electricity would be valued at 141 €/MWh. More favourable conditions could reduce the gap considerably (see section 6.3). A price reduction of the SOFC system alone does not suffice to make the system profitable. The price of the two stage gasification and gas cleaning has to decrease too. For break-even at 70 €/MWh, prices would have to drop e.g. from 1600 €/kW$_e$ to 500 €/kW$_e$ for the SOFC and from 200 €/kW$_{th}$ to 80 €/kW$_{th}$ for the gasification and gas cleaning.

In conclusion, it is fair to say that the present equipment prices would allow profitable exploitation only in very specific favourable circumstances.

6.2 2.5 MW Green MoDem

Green MoDem is the name given by TechnoInvent to a system it has designed for decentralized conversion of high-caloric waste into power and heat. The Green MoDem uses a proprietary technique to convert waste into oil and gas. ECN has not been involved in the development of the technique or in the design of the Green MoDem. The system and economic analysis of the Green MoDem have been performed independently by TechnoInvent. ECN has only provided data on the composition of waste fuel and on the expected SOFC performance and costs.

A full description and analysis of the Green MoDem by TechnoInvent can be found in Appendix A (in Dutch). Here, only a summary is given.

6.2.1 System description

The Green Modem uses two reactors to convert waste fuel into oil and gas (see Figure 6.2). The first reactor produces oil, gas and char by thermal cracking. Char is transported to the second reactor and gasified by contact with the hot combustion products of the gas and part of the oil produced in the first reactor.

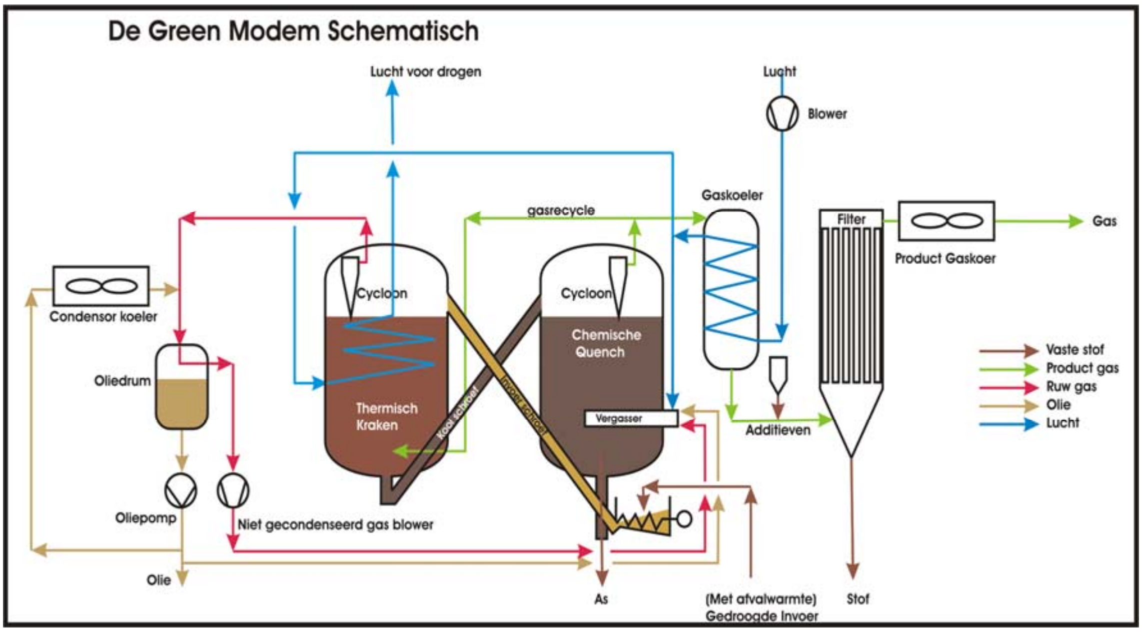

Figure 6.2 *Schematic representation of the Green MoDem process for the production of oil and gas from waste fuel.*

The economic analysis has been performed for three cases. The first or base case corresponds to a system as marketed by TechnoInvent. Oil and gas are fired in a dual-fuel engine. The second case is a variant in which the gas is fed to an SOFC system. Depleted off-gas and oil are fired in a dual-fuel engine. In both cases, locally available industrial waste fuel is used at an existing site. The third case is identical to the second case, but applies to a newly built site and use of pre-treated waste fuel delivered by a waste collector.

6.2.2 System performance

Performance data of the Green MoDem systems for the base case and the two SOFC cases are given in Table 6.4. The fuel considered has a composition similar to Rofire (see Table 3.1) but a moisture content of 25%. The systems include a fuel drier. Heat loss is neglected.

6.2.3 System economy

Table 6.5 gives the investment costs for Green MoDem systems. The total cost of equipment has been multiplied by a factor 4 to account for installation etc. on an existing site. A factor 4.5 is used in scenario 2, which applies to a plant built at a new site. No multipliers have been applied to the costs of the dual fuel engine (300 €/kW$_e$), SOFC system and other equipment that is delivered ready for use. The shredder is needed only in the base case and in SOFC scenario 1 for pre-treatment of the locally available industrial waste fuel. The total specific investment amounts to 2800 €/kW$_e$ for the base case and to approximately 4000 €/kW$_e$ for the systems with SOFC and dual fuel engine.

Table 6.4 *Performance data of Green MoDem (GM) systems.*

	GM+Dual Fuel engine	*GM+SOFC*	*unit*
Capacity	0.53	0.53	t/h
Thermal capacity input	2.61	2.61	MW (LHV)
Oil production	1.36	0.53	MW (LHV)
Gas production	1.18	1.97	MW (LHV)
E consumption	0.06	0.13	MW
E production	1.10	1.30	MW
E export	1.04	1.18	MW
Heat export	1.07	0.39	MW
GM thermal efficiency	97.4	95.8	%
E efficiency	39.6	45.1	%
Heat efficiency	41.0	15.0	%
System total efficiency	80.6	60.1	%

Table 6.5 *Investment costs of Green MoDem systems.*

		Base case		*SOFC scenario 1*		*SOFC scenario 2*
Investment SOFC infra	€	0	€	115,301	€	115,311
Investment bunker	€	69,131	€	69,132	€	69,132
Total installations	€	69,131	€	184,433	€	184,443
Multiplier		4.00		4.00		4.50
Investment infrastructure	€	276,523	€	737,733	€	829,992
Investment Green Modem	€	2,000,000	€	2,000,000	€	2,000,000
Investment shredder	€	90,000	€	90,000	€	0
Investment mobile	€	15,000	€	15,000	€	15,000
Investment E generation	€	330,000	€	1,540,000	€	1,540,000
Working capital	€	200,000	€	200,000	€	200,000
Total investment	€	2,911,523	€	4,582,733	€	4,584,992

The assumptions used in the economic analysis are given in Table 6.6. There are several differences with the assumptions used in the previous Section (c.f. Table 6.1). The main differences are 100% vs. 50% use of available heat, an interest rate of 8% instead of 10%, 2.5% insurance & overhead plus 25% tax on profit instead of 2% insurance & tax, operator costs of 35 k€/person.yr instead of 60 k€/person.yr, and fuel prices of -100 €/tonne or 0 €/tonne instead of -20 €/tonne. The Green MoDem analysis also includes CO_2 credits in the base case and scenario 1, and subsidy for electricity delivered to the grid in scenario 2.

The economic analysis leads to a simple pay-back time of 4.9 years for the base case. The SOFC scenarios have a pay-back time of 10.5 and 5.7 years. The SOFC scenarios are more sensitive to changes in the CO_2 credits and subsidy than the base case. The higher electrical efficiency of SOFC systems does not suffice to justify investment at a price of 1600 €/kW at present electricity prices.

Table 6.6 *Basic assumptions for economic analysis of Green MoDem*

Economic parameter	*Value*	*Unit*
Power plant life	15	Year
Load factor	85	% (7500 hr/yr)
Heat utilization factor	100	% (7500 hr/yr)
Interest	8	%/year
Insurance & overhead	2.5	%/year of plant investment including electric generators
Maintenance costs	4	%/year of plant investment excluding electric generators 15 €/MWhr net production for electric generators [1]
Number of operators	0.3 / 1.2	0.3 base case + scenario 1, 1.2 in scenario 2
Operator cost	35	k€/yr
Fuel (biomass) price	-100 / 0	-100 €/tonne base case + scenario 1, 0 €/tonne delivered at site in scenario 2
Gas price	7	€/GJ (= 25 €/MWh)
Electricity (market) price	70	€/MWh
CO_2 credits	10	€/tonne
Adsorbent price	200	€/tonne
Water price	1	€/m^3
Ashes disposal price	150	€/ton
Subsidy	0 / 96	€/MWh, 0 in base case + scenario 1, 96 in scenario 2

[1] The SOFC system requires comparatively little maintenance, but the stack will have to be replaced after 40,000 to 80,000 hours of operation.

6.3 Discussion

The economic analyses of the 25 MW and 2.6 MW systems reported above indicate that in the near future SOFC systems will not be economically viable or competitive at the expected price level of 1600 €/kW_e. The results also suggest that the Green MoDem is superior to the ECN two-stage gasification system. However, the assumptions made for the economic analyses by ECN and TechnoInvent are too different to allow a direct comparison. To that end, Table 6.7 shows the results for the ECN system for both sets of assumptions. The simple pay-back time would reduce from 36 years to 9.3 years. That value should be compared to the 10.5 years for the similar SOFC scenario 1 for the Green MoDem. At 10% interest, the ECN system would become profitable from an electricity price of 87 €/MWh. At 8% interest and 70 €/MWh, the system would become profitable at an SOFC price of 1250€/kW_e.

Table 6.7 *Economic analysis of ECN two-stage gasification system with SOFC and steam bottoming cycle for different economic parameters.*

	Case 1vs Case 2[#]	*Case 1*	*Case 2*
Revenues (M€/yr)		**6.45**	**10.20**
Electricity	-	5.01	5.01
Heat	50% vs 100%	0.86	1.72
Fuel	-20 vs -100 €/tonne	0.58	2.89
CO_2 credits	0 vs 10 €/tonne	-	0.58
Variable costs		**5.09**	**4.95**
Maintenance	-	2.08	2.08
Insurance (& overhead)	2% vs. 2.5%	0.82	1.02
Operator costs	60 vs 35 k€/person.yr	0.90	0.53
Oxygen	-	0.87	0.87
Adsorbents + water	130 vs 200 €/tonne; 1.31 vs 1 €/m^3	0.08	0.11
Ash	-	0.34	0.34
Gross profit	-	**1.36**	**5.25**
Simple pay-back time (years)		**36**	**9.3**

[#] Parameters given in Table 6.1 and Table 6.6 respectively, number of operators from Table 6.1.

7. Conclusions

Very encouraging experimental results have been obtained:

- Soot formation in the second (high temperature) stage of the gasification process can be suppressed effectively by steam addition.
- Soot formation depends on the type of fuel used. Beech wood produces much less soot than Rofire pellets which contain a substantial amount of plastics.
- Commercial adsorbents for S and Cl can be used at 200°C to 300°C to clean syngas sufficiently for use in an SOFC system.
- The commercial SOFC stack used showed no degradation during tests on syngas from three different fuels for 250 hours in total.

Long-term operation of the commercial SOFC stack on a mixture of H_2 and N_2 (not part of this project) showed a performance degradation of less than 1% per 1000 hours over a period of more than 4500 hours [4]. According to the manufacturer, stack quality is continuously improved and has reached a level where stacks show only 0.4% degradation per 1000 hours. Part of the work described in this report has been presented at the 16th European Biomass Conference & Exhibition in 2008 [10].

The system analysis led to the following conclusions:

- Two-stage gasification of dry waste fuel with syngas use in an SOFC system would yield a net electrical efficiency of only 17% if part of the syngas were burned in the second gasification step.
- The net electrical efficiency could be increased to 22% if SOFC fuel utilization could be increased from 60% to 80%.
- High SOFC fuel utilization would probably be incompatible with high-temperature syngas cleaning.
- Addition of a steam bottoming cycle could increase the net electrical efficiency to 27%.
- Steam gasification of char produced in the first gasification stage could increase the net electrical efficiency to 37%.
- Reduction of the auxiliary power consumption could bring the net electrical efficiency close to 40%.

The economical analysis led to the following conclusions:

- A 25 MW_{th} system with two-stage gasification, SOFC and steam bottoming cycle would not be economically viable at the assumed conditions. Break-even could be reached if electricity would sell at 141 €/MWh.
- Both the SOFC system and the two-stage gasification and gas cleaning have to become cheaper. For break-even at 70 €/MWh, prices should decrease e.g. from 1600 €/kW_e to 500 €/kW_e and from 200 €/kW_{th} to 80 €/kW_{th}.
- More favourable conditions could reduce the single pay-back time of the system to 9.3 years.
- With these favourable conditions and 10% interest, break-even would be reached for an electricity price of 87 €/MWh.
- With these favourable conditions and 8% interest rate, break-even would be reached for an electricity price of 70 €/MWh if the SOFC price would decrease to 1250 €/kW_e.
- A 2.6 MW_{th} Green MoDem system which uses a different two-stage gasification technique could reach nearly 40% net electrical efficiency plus 41% thermal efficiency from waste fuel with 25% moisture.
- With the same more favourable economic conditions mentioned above, the Green MoDem system with dual-fuel engine would have a single pay-back time of 4.9 years.
- A modified Green MoDem system with SOFC system and dual-fuel engine would have a single pay-back time of 10.5 years for the same conditions.

- A modified Green MoDem system with SOFC system and dual-fuel engine would have a single pay-back time of 5.7 years if subsidy for renewable electricity production could be obtained.

For both systems, the high electrical efficiency of SOFC systems does not suffice to pay the price expected in the near future for SOFC systems. On the other hand, even the largest SOFC system in development will be a factor 4 smaller than needed for the Green MoDem system. Some development at that front will be needed too. If in the mean time the electricity price will keep rising, and if the goals of cost reduction for SOFC systems will be met, there may still be a future for two-stage gasification combined with SOFC systems.

8. References

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[2] Tar protocol, see www.tarweb.net

[3] *SOFC / Gas Turbine Hybrid*. Siemens AG 2007, corporate information, see www.powergeneration.siemens.com/products-solutions-services/products-packages/fuel-cells

[4] *Operation of a Staxera SOFC stack fuelled with cleaned gas from a gasifier*. N. Dekker, B. van der Drift, L. Rabou, H. van Wees and B. Rietveld. Contribution B1104 to the 8th European SOFC Forum, Luzern, 2008.

[5] Press release 080124 and flyer 080408, see www.staxerafuelcells.de.

[6] Press release June 2008, see www.topsoefuelcell.com.

[7] *Conceptual study of a 250 kW planar SOFC system for CHP application*. E. Fontell, T. Kivisaari, N. Christiansen, J.-B. Hansen and J. Pålsson. Journal of power Sources 131 (2002) 49-56.

[8] *Onrendabele top berekeningen voor nieuw WKK vermogen 2008.* J.S. Hers, W. Wetzels, A.J. Seebregts, A.J. van der Welle. Report ECN-E--08-016.

[9] *Conversietechnologieën voor de productie van elektriciteit en warmte uit biomassa en afval, EWAB Marsroutes Taak 2.* J.A. Zeevalkink and R. van Ree. Report 2EWAB00.22.

[10] *Fuel cell stack operation on clean fuel gas from rejects of paper recycling*. L.P.L.M. Rabou, N.J.J. Dekker and B. Vreugdenhil. Contribution W2916 to the 16th European Biomass Conference & Exhibition, Valencia, 2008.

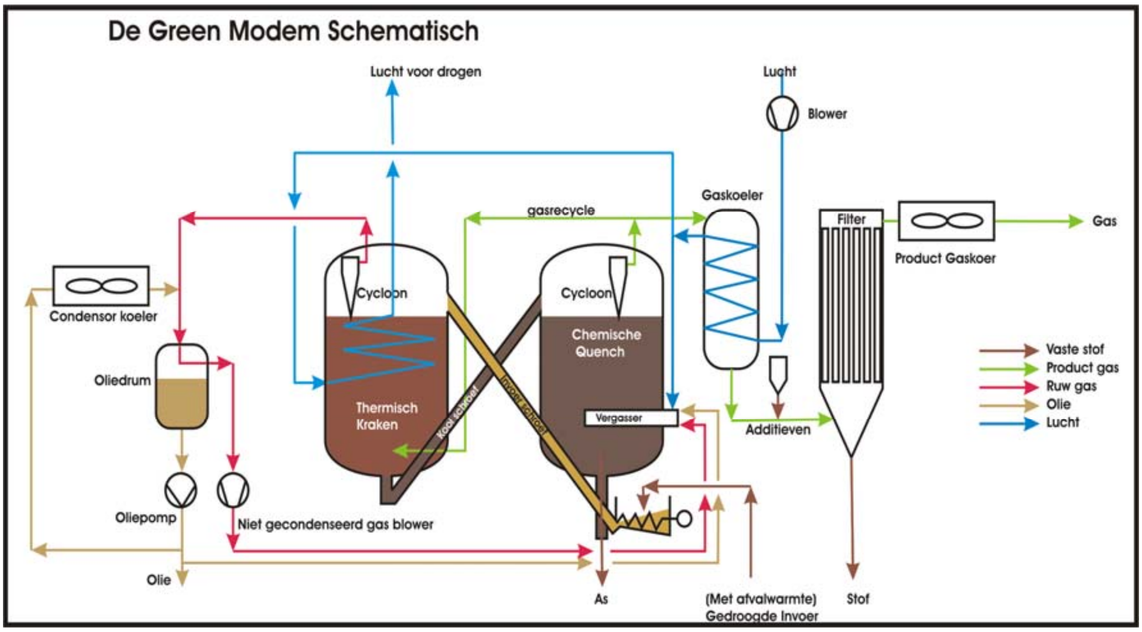
Appendix A Green MoDem

Het proces is specifiek ontwikkeld voor decentrale behandeling van biomassa en hoogcalorische afvalcomponenten. De naam Green MoDEM is gekozen in analogie met het internet: de Green MoDEM verbindt decentraal opgewekte energie en gebruikers via het (elektriciteitskabel) netwerk.

De capaciteit bedraagt, afhankelijk van het type invoer 2-3 kton droge stof per jaar. Het proces is ingebouwd in een gemakkelijk over de weg te transporteren staalframe. De Green MoDem wordt op de markt gebracht door Green Energy Technologies, een samenwerking tussen TechnoInvent en RL Finances. Zie ook www//get-technologies.com

Procesbeschrijving

Het proces omvat 4 achtereen geschakelde stappen: Drogen met restwarmte, Thermisch kraken, Vergassing en Chemische quench.



Drogen met restwarmte (niet in de figuur)
De invoer wordt indirect gedroogd met een circulerende gasstroom in een gesloten systeem. De circulerende gasstroom bestaat bijna uitsluitend uit waterdamp. Daardoor kan bij condensatie heet water tot 90°C worden opgewekt.

Thermisch kraken
De thermische kraakstap wordt uitgevoerd in een fluïde bed. Zand of Katalysator is het bedmateriaal. De invoer wordt gekraakt in kool, olie en gas. Door de juiste keuze van de condities ontstaat een dampmengsel van niet met water mengbare, neutrale olie en lichte componenten. De verblijftijd van de damp bedraagt enige seconden. Het gasmengsel wordt gekoeld tot 120-150°C. Olie condenseert en wordt afgescheiden. Water en lichte producten blijven gasvormig.

Vergassing

De vergassingsreactor bestaat uit een brander en een reactiekamer. Het niet gecondenseerde gas en een deel van de olie worden met voorverhitte lucht vergast boven 1200°C (partiële oxidatie). De reactie verloopt binnen een tiende seconde onder vorming van een mengsel van N_2, CO, H_2 en CO_2. Halogeen, zwavel en stikstof verbindingen tot resp. HF, HBr, HCl, H_2S en N_2.

Chemische quench

De Chemische quench wordt uitgevoerd in een fluïde bed, gevuld met kool en het opgekoolde bedmateriaal. In de chemische quench wordt het hete afgas van de vergassing gekoeld door de endotherme reactie van CO_2 en water met kool. Daarbij wordt extra CO en H_2 gevormd.

De chemische quench draait bij een temperatuur tussen 750-850°C (ver onder de temperatuur waar asdelen gaan sinteren en zouten vervluchtigen). Geregenereerd bedmateriaal wordt teruggebracht in de thermische kraakreactor.

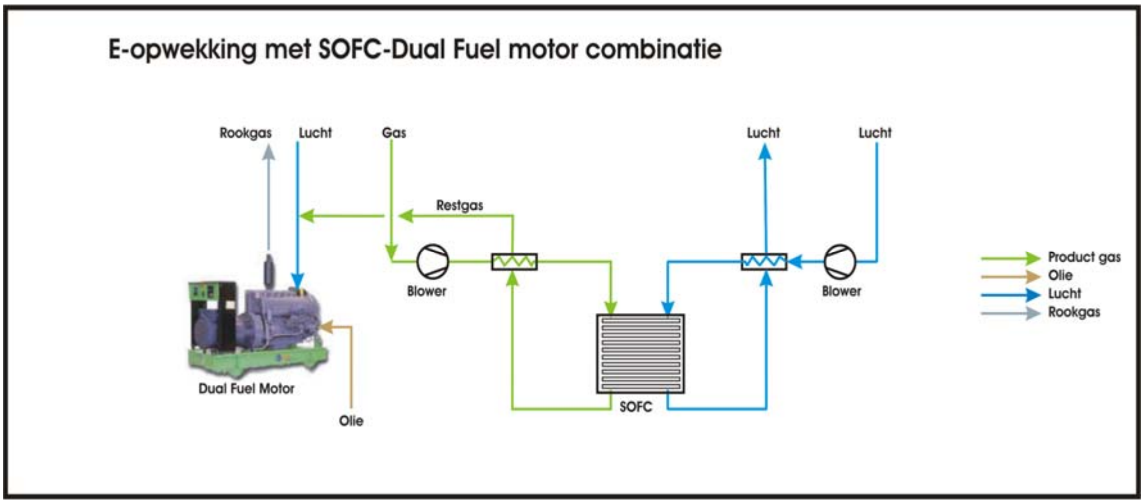
Gasreiniging

Het hete gas wordt gekoeld tegen lucht. HF, HBr en HCl worden gebonden met een adsorbent. Stof en roet worden verwijderd in een filter. Na het filter wordt Zwavel verwijderd aan ZnO. Bij verder koelen vormt zich schoon condensaat. Het gas bevat 50-60% CO+H_2. Omdat het gas geen teer bevat en de verontreinigingen gebonden worden kan het gefilterde gas worden ingezet in een SOFC.

Productie van elektriciteit en afvalwarmte

Elektriciteit wordt geproduceerd in een SOFC en een dual fuel motor. De dual-fuel motor wordt bedreven met spuigas uit de SOFC en de rest van de geproduceerde olie. Lucht en gas worden voorverwarmd tegen de effluentstromen uit de SOFC.

Afvalwarmte (inclusief) de condensatie warmte uit de droger wordt opgenomen in een heet watercircuit met een aflevertemperatuur van 90°C.



Toepassing van de Green MoDem

De Green MoDem wordt volledig gemonteerd in een staalframe, 2.5 x 12 m, 3 m hoog, De Green MoDem kan op een stevige ondergrond (bij voorkeur stelcon platen) worden geplaatst. Het totale volgewicht bedraagt ca 50 ton.

De Green MoDem kan snel in bedrijf komen: Levertijd, inbedrijfname en opstarten bedragen 8-12 maanden.

De Green MoDem is verregaand geautomatiseerd en kan op afstand worden bediend. Voor de logistiek en toezicht is personeel in deeltijd nodig.
De Green MoDem is daarom bijzonder geschikt als een utility-eenheid bij een of meer industrieën voor het omzetten van het eigen afval in elektriciteit en warmte.
De Green MoDem kan de brandbare fractie van mechanisch gescheiden afval van ca 5000 gezinnen omzetten in elektriciteit, voldoende voor ca. 2000 gezinnen.

Uitgangspunten voor het ontwerp

Uitvoering
De installatie is ontworpen voor het omzetten van 3000 ton droge stof per jaar gedurende 7500 draaiuren.
In het stalen frame zijn de schakelkast, de droger, de reactoren, het filter, de blowers en pompen en de warmtewisselaars, inclusief de warmtewisselaars voor het verwarmen van de lucht en het gas voor invoer in de SOFC opgenomen. De Dual Fuel motor, de SOFC en de olieopslag zijn in een apart frame ondergebracht.

Werkwijze
Het ontwerp bestaat uit het opstellen van de massa-en warmtebalansen, het kiezen van de apparaten en het bepalen van de specificaties, het selecteren van de materialen en het bepalen van de wijze van bediening en automatisering.
De ontwerp-berekeningen werden uitgevoerd met een computermodel waarin massa-warmtebalansen worden berekend aan de hand van samenstelling van de invoerstromen met thermodynamische evenwichten en chemische kinetiek. Tevens worden in het model de apparaten en leidingdiameters berekend, de elektriciteitsbehoefte en de productie van elektriciteit. Tenslotte bevat het model een module voor het schatten van de investeringskosten en het uitvoeren van economische evaluaties.
Het model is zeer geschikt voor het berekenen van recycle stromen, en convergeert in een zeer groot bedrijfsvenster.

Het model is geverifieerd aan tal van tijdens proefruns verkregen testresultaten van PKA, enkele door ECN uitgevoerde testen, de met zuurstof bedreven installaties van Oxytech, een smelter/vergasser in Velmede en de eigen testinstallatie. Dit geldt ook voor de samenstelling van de producten. Het model werkt uitstekend binnen een werkgebied van omgevingstemperatuur tot en met ca 2000°C.

Een eigen testinstallatie is gebouwd en bedreven voor het verifiëren van de berekeningsresultaten en het bepalen van de meest geschikte bedrijfscondities. Uit de onder gegeven vergelijking tussen de berekende samenstelling van de olie en de analyseresultaten blijkt de accuratesse van het model.

Ontwerpoverwegingen
Omdat het afgas van de droger praktisch geheel uit waterdamp bestaat, kan daaruit bij condensatie heet water worden geproduceerd. Om ook hele vochtige invoerstromen aan te kunnen, wordt de droger bedreven met heet afgas van de motor.

Door toepassing van de chemische quench en het toepassen van een met restwarmte (het rookgas van de dual-fuel motor) heeft de Green MoDem een hoge thermische efficiency: voor de hier ingezette invoer ca 95%. Hierbij zijn warmteverliezen (geschat op enkele procenten) niet meegenomen.

De temperaturen in de reactoren zijn zo gekozen, dat geen keramische materialen nodig zijn, alle reactoren zijn vervaardigd uit (hoogwaardig) roestvrij staal. Voor de hoge temperatuur toepassingen is een speciale legering gekozen.

Lokaal (bij de uitlaat van de brander) kan de temperatuur in de chemische quench reactor boven het verwekingpunt van de as komen. Daardoor kunnen agglomeraten worden gevormd. De reactor is daarom uitgerust met een ruim bemeten asafvoer.
De combinatie van de SOFC en de Dual Fuel motor heeft grote voordelen. Spuigas wordt alsnog met ca 40% rendement omgezet in elektriciteit, zodat het E-rendement hoog uitkomt.

Nadeel van het toepassen van een SOFC is het relatief hoge E-verbruik voor het bedrijven van de blowers voor lucht en gas. Daarom is gekozen voor een systeem met lage drukval, dwz relatief korte, brede warmtewisselaars en grote diameter, korte pijpen zijn daarom gekozen. Afgezien is van het gebruik van een heetgas circulatie. De daarmee te sparen elektriciteit weegt niet op tegen de bedrijfskosten van een hete lucht gas circulatie, zowel een (uit exotisch materiaal vervaardigde recycleventilator als een met perslucht bedreven eductor zijn daarbij onderzocht.

Gekozen is voor een gasreiniging bij een temperatuur boven 200°C. Daardoor kan worden voorkomen dat zure componenten in de pijpen en het filter condenseren. Pas na het verwijderen van de zure componenten wordt het gas verder gekoeld en kan het op de gewenste druk worden gebracht. Het condensaat dat hier wordt gevormd is relatief schoon. Omdat het in het filter niet verwijderd stof dient als condensatiekern, wordt het gas in de nakoeling nog verder ontstoft.

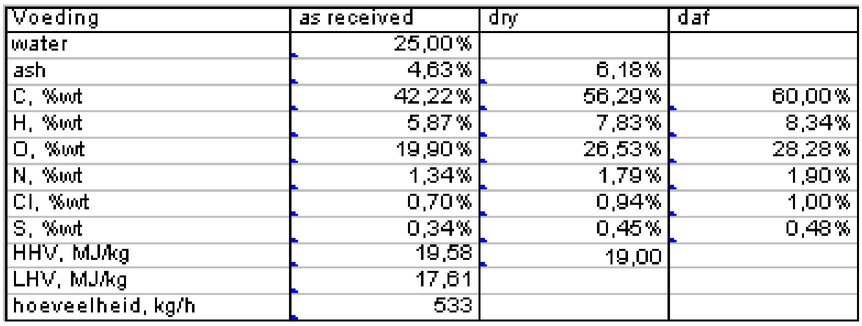
De invoerstromen

Het concept van de Green MoDem is het decentraal produceren van hoogwaardige energiedragers uit decentraal beschikbare brandstoffen zoals Rejects uit de papierindustrie, RDF, tapijtafval, banden, afgewerkte olie, hout etc. In feite is de Green MoDem een *raffinaderij* voor vaste stof.

Decentrale omzetting naar hoogwaardige energiedragers bespaart veel transport, maar verruimt ook de mogelijkheid restwarmte af te zetten. Daardoor kan veel worden bespaard op de kosten voor afvoer en verwerking van afval.

Afval bestaat uit een mengsel van biomassa, kunststoffen, metalen en mineralen. Met mechanische voorbewerking kan afval worden gescheiden in een mineralen, metalen en een laagcalorische en een hoogcalorische fractie.
Al deze stromen kunnen worden behandeld, mits de deeltjesgrote beperkt is tot enkele cm's. Extreem natte (>50% vocht) of zeer asrijke stromen zijn minder geschikt omdat daarmee een slecht thermisch rendement wordt behaald.

In het algemeen is mechanische voorscheiding en verkleining gewenst. De daarvoor in te zetten technieken zijn goed bewezen en worden in toenemende mate toegepast. Momenteel worden hoogcalorische fracties verbrand voor opwekking van energie en deels ook gestort. De laagcalorische fracties worden gestort of biologisch gedroogd en nagescheiden.
Voor de studie zijn de afvalstromen beschouwd, die vrijkomen op de locatie waar de eerste Green MoDem eind 2008 zal worden geïnstalleerd. Een representatief monster is door ECN geanalyseerd. De voor het ontwerp gebruikte samenstelling en de analyseresultaten zijn hieronder gegeven.

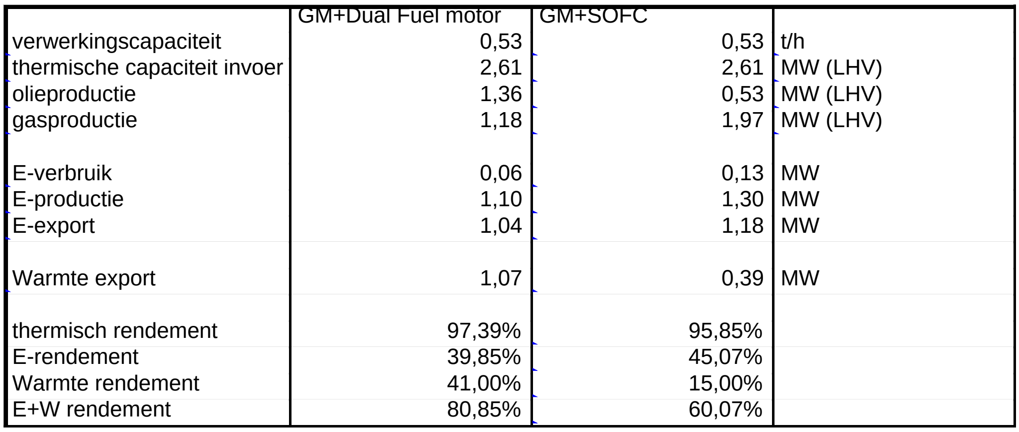


Voeding	as received	dry	daf
water	25,00%		
ash	4,63%	6,18%	
C, %wt	42,22%	56,29%	60,00%
H, %wt	5,87%	7,83%	8,34%
O, %wt	19,90%	26,53%	28,28%
N, %wt	1,34%	1,79%	1,90%
Cl, %wt	0,70%	0,94%	1,00%
S, %wt	0,34%	0,45%	0,48%
HHV, MJ/kg	19,58	19,00	
LHV, MJ/kg	17,61		
hoeveelheid, kg/h	533		

Energiebalans

De energie balansen zijn berekend voor de systemen:

- Green MoDem in de bovenbeschreven combinatie met een SOFC (spuigas in een Dual fuel motor)
- Green MoDem met een dual fuel motor



	GM+Dual Fuel motor	GM+SOFC	
verwerkingscapaciteit	0,53	0,53	t/h
thermische capaciteit invoer	2,61	2,61	MW (LHV)
olieproductie	1,36	0,53	MW (LHV)
gasproductie	1,18	1,97	MW (LHV)
E-verbruik	0,06	0,13	MW
E-productie	1,10	1,30	MW
E-export	1,04	1,18	MW
Warmte export	1,07	0,39	MW
thermisch rendement	97,39%	95,85%	
E-rendement	39,85%	45,07%	
Warmte rendement	41,00%	15,00%	
E+W rendement	80,85%	60,07%	

Uit de berekeningsresultaten komen de sterke en de zwakke kanten van het toepassen van de SOFC naar voren:

STERK: Het overall elektrisch rendement ligt aanzienlijk hoger dan bij toepassing van een Dual fuel motor. Netto extra opbrengst ca 1000 MWh/jaar

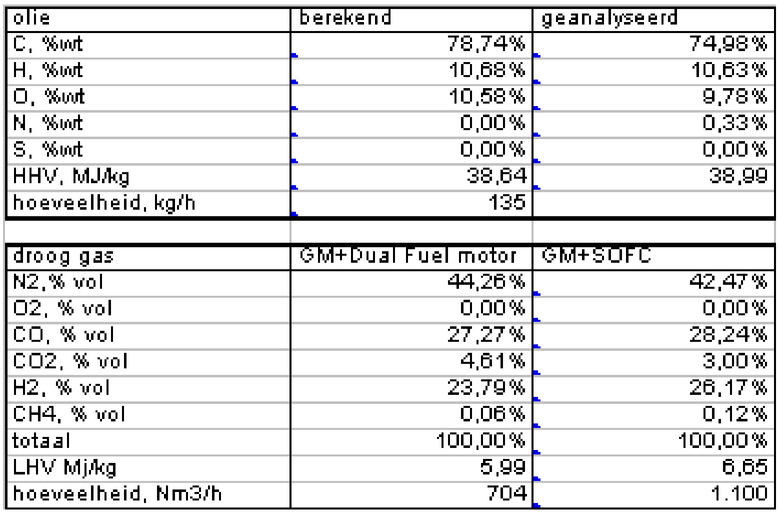
ZWAK: Doordat een groot deel van de olie wordt vergast daalt het thermisch rendement (vergassen van olie is endotherm)

Het eigen E-verbruik is meer dan twee maal zo groot (extra energiebehoefte voor het comprimeren van de lucht naar de SOFC)

Een laag warmte rendement: omdat lucht naar de SOFC moet worden voorverwarmd komt de restwarmte uit de SOFC bij een te laag niveau vrij.

De producten

De Green MoDem produceert geen rookgas maar brandbaar gas en olie. Rookgas ontstaat bij de omzetting naar elektriciteit. De samenstelling van de olie en het gas zijn berekend voor beide bovengenoemde systemen. In een proefinstallatie geproduceerde olie is door ECN geanalyseerd. Een door TU Eindhoven uitgevoerde GCMS analyse is in de bijlage gegeven.



olie	berekend	geanalyseerd
C, %wt	78,74%	74,98%
H, %wt	10,68%	10,63%
O, %wt	10,58%	9,78%
N, %wt	0,00%	0,33%
S, %wt	0,00%	0,00%
HHV, MJ/kg	38,64	38,99
hoeveelheid, kg/h	135	

droog gas	GM+Dual Fuel motor	GM+SOFC
N2,% vol	44,26%	42,47%
O2, % vol	0,00%	0,00%
CO, % vol	27,27%	28,24%
CO2, % vol	4,61%	3,00%
H2, % vol	23,79%	26,17%
CH4, % vol	0,06%	0,12%
totaal	100,00%	100,00%
LHV Mj/kg	5,99	6,65
hoeveelheid, Nm3/h	704	1.100

De berekende samenstelling van de olie blijkt weinig af te wijken van de analysewaarde. Dit is een gevolg van de toegepaste berekeningsmethodiek: daarbij wordt uitgegaan van een combinatie de stofkarakterisering, gemeten omzettingskinetiek en ervaringsgetallen.

In de invoer aanwezige niet brandbare componenten worden gesplitst in een grove (> 80 µm) en fijne fractie. In het algemeen kan de grove fractie worden hergebruikt. In de fijne fractie zijn in de invoer aanwezige zware metalen en in de gasreiniging ontstane zouten (vooral Chloriden) geconcentreerd. In het algemeen moet de fijne fractie daarom worden gestort.

Elektriciteit wordt omgevormd naar 50Hz 400V (3 fasen).
Warmte wordt geleverd als heet water, inlaat 60 °C, uitlaat 90°C.

Gasreiniging
De gasreiniging is ontworpen conform de EC-directive. Door toepassing van een met ZnO gevulde guardreactor worden H_2S en halogenen onder 0,5 ppm gehouden. De Dual Fuel motor is uitgerust met een SCR.

DIRECTIVE 2000/76/EC	**Verwachte emissie**	**EC directive***
Total dust, mg/m^3	<1	30
HCI , mg/m^3	<0,1	10
HF, mg/m^3	<0,1	1
NO_x , mg/m^3	<500	500
Hg , mg/m^3	0,01	0,05
Cd + Tl , mg/m^3	0,01	0,05
Heavy metals, mg/m^3	0,1	0,5
Dioxins and furans ng/m^3	0,05	0,1
SO_2 , mg/m^3	<50	50
TOC , mg/m^3	<10	10

Economische evaluatie

De Economische evaluatie omvat de volgende onderdelen:

- Kiezen van de te onderzoeken scenario's
- Schatten van de investering
- Vast stellen van de variabele kosten/opbrengsten (grond- en hulpstoffen, producten en reststromen)
- Vaststellen van de vaste laste lasten (personeel, onderhoud, verzekeringen en belastingen).
- Vaststellen van de financiële condities (rente, aflossingstermijnen, restwaarde)
- Uitvoeren van de evaluatie

De scenario's
De scenario's hebben een grote invloed op de investering, de kosten en de opbrengsten. Daarom is gekozen voor twee "uitersten". De scenario's zijn vergeleken met de "base case".

Base case: De Green Modem verwerkt op het terrein van een industrie daar ontstane afvalstoffen en zet elektriciteit en restwarmte af aan deze industrie. Daarmee bespaart de industrie aan afvalkosten (tot meer dan 100 €/per ton, en inkoop van elektriciteit. Gebruik van restwarmte bespaart aardgas. Bestaand personeel kan (in deeltijd) worden ingezet voor bediening en logistiek. Door de beschikbaarheid van bestaande infrastructuur (wegen, gebouwen, terreinafscheiding, voorzieningen (elektriciteitskabels en trafo's, gas, perslucht, proces- en bluswater) kan worden gerekend met een lagere opstelfactor dan voor een "greenfield" installatie. Daartegenover staat extra investering voor mechanische voorbewerking en opslag. De olie wordt niet vergast maar samen met het productgas omgezet in elektriciteit in een dual fuel motor, geen SOFC is geïnstalleerd.

Scenario 1: Opzet als de base case met uitzondering van de elektriciteitsproductie. Ca 65% van de olie wordt vergast, het gas wordt omgezet in elektriciteit in een SOFC, het spuigas van de SOFC wordt samen met de rest van de olie omgezet in elektriciteit in een dual fuel motor.

Scenario 2: De Green Modem produceert warmte en elektriciteit op een geschikte plaats, bijvoorbeeld bij kassen of een woonwijk met districtverwarming. Elektriciteit en warmte worden via nieuwe netaansluitingen afgezet. Grondstof wordt betrokken van een afvalbewerker. Voor bediening en logistiek is t.o.v Scenario 1 FTE extra personeel nodig. Elektriciteitsproductie als in scenario 1.

De investering

Voor de raming van de investering uit uitgegaan van:

1. De Green Modem compleet € 2.000.000.
2. Voor de installatiedelen: opslag van invoer en producten en de infrastructuur rond de SOFC) zijn de prijzen van de apparaten bij leveranciers opgevraagd en vermenigvuldigd met een opstelfactor. Deze factor omvat de kosten voor staal, pijpen en appendages, instrumentatie, elektrische voorzieningen en bekabeling, civieltechnische voorzieningen, montage, engineering, in bedrijf stellen en de posten onvoorzien en winst. De hoogte van de opstelfactor is afhankelijk van de mogelijkheid, bestaande infrastructuur te gebruiken, dus van het gehanteerde scenario.
3. Voor de shredder en het intern transport zijn de prijzen van de apparaten bij leveranciers opgevraagd, geen opstelfactor is hier gehanteerd, omdat het hier gaat om complete, mobiele eenheden.
4. Voor de Elektriciteitsopwekking (dual-fuel motor en SOFC) zijn specifieke investeringskosten gehanteerd, uitgedrukt in €/kW elektrisch vermogen. Deze kosten omvatten de apparatuur, elektrische aansluitingen, voorzieningen zoals geluidskasten en koeling, regelapparatuur en schakelkast.

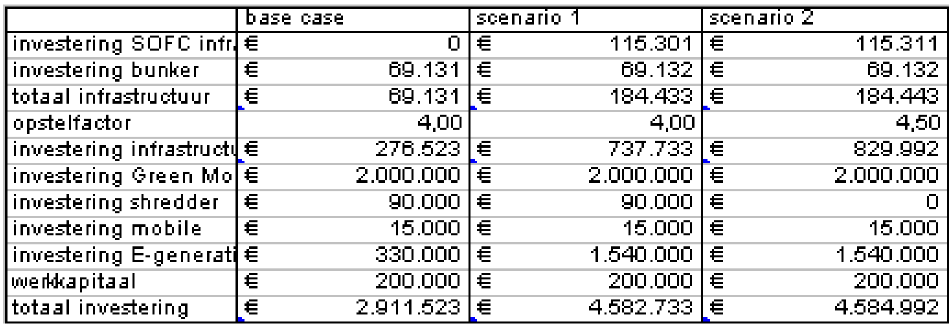
Leveringsomvang:

- Green Modem: Alle reactoren, invoersysteem, vaten, warmtewisselaars, inclusief noodkoeling en heet water systeem, pompen, compressoren, doseersystemen, filters, branders, noodfakkel. De Green MoDem wordt compleet gemonteerd en getest in een staalframe afgeleverd. Voor de fundatie volstaat een vlakke vloer.
- Invoervoorbereiding en opslag: overdekte opslag voor invoer (1 week), containers voor as, stof en olie, begroot op € 78.900 en een shredder/zeef (scenario 1), begroot op € 90.000. Het electriciteitsgebruik van de shredder is gesteld op 12 kWh/ton invoer (opgave leverancier).

- Infrastructuur SOFC: de warmtewisselaars en de blowers voor lucht en gas, begroot op € 113.500.
- SOFC: vooralsnog (zolang de stacks nog relatief duur zijn en onbeproefd voor langere bedrijfstijden) is serieproductie geen optie. De SOFC is daarom begroot op de 1500 €/kW. Dit is de verwachte prijs, als serieproductie op gang zal zijn gekomen. Voor het omzetten van de opgewekte stroom naar 400V wisselstroom is een post van 165 €/kW opgenomen.
- Dual fuel motor: In tegenstelling tot een gasmotor is de kwaliteit van het ingevoerde gas niet van invloed op het vermogen van de motor. De hoeveelheid in te voeren gas wordt uiteraard wel beperkt door de kans op pingelen, dus de explosiegrens van het mengsel van verbrandingslucht en het ingevoerde gas. Daarom wordt de motor begroot op 300 €/kW.

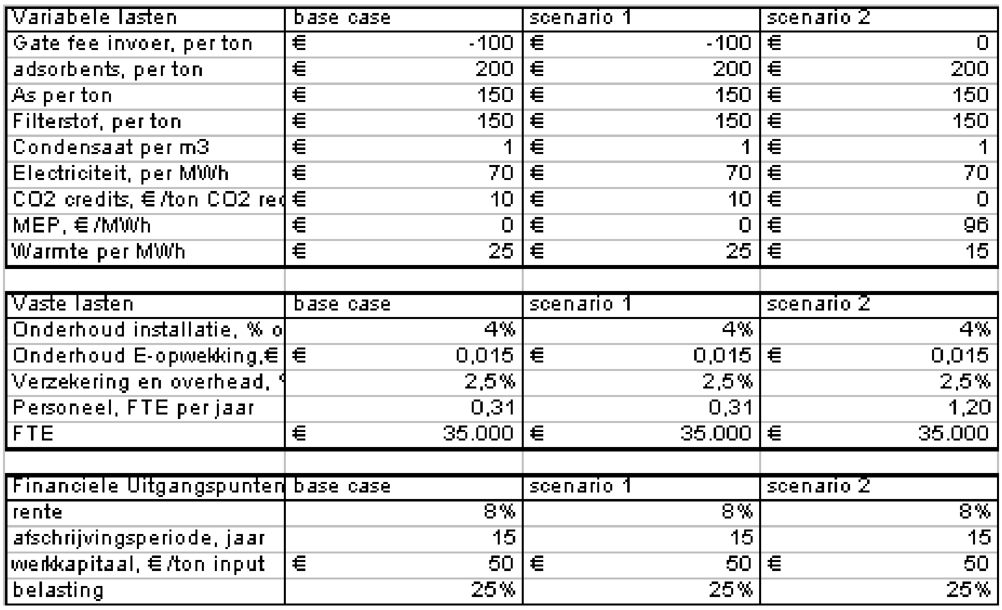
Opstelfactoren:
Voor de base case en scenario 1 (inpassing in bestaande infrastructuur) is een opstelfactor voor de infrastructuur van 4.0 gehanteerd, voor scenario 2 een opstelfactor van 4,5.

De investeringsramingen voor de base case en de twee scenario's zijn hieronder gegeven.



	base case	scenario 1	scenario 2
investering SOFC infr	€ 0	€ 115.301	€ 115.311
investering bunker	€ 69.131	€ 69.132	€ 69.132
totaal infrastructuur	€ 69.131	€ 184.433	€ 184.443
opstelfactor	4,00	4,00	4,50
investering infrastructu	€ 276.523	€ 737.733	€ 829.992
investering Green Mo	€ 2.000.000	€ 2.000.000	€ 2.000.000
investering shredder	€ 90.000	€ 90.000	€ 0
investering mobile	€ 15.000	€ 15.000	€ 15.000
investering E-generati	€ 330.000	€ 1.540.000	€ 1.540.000
werkkapitaal	€ 200.000	€ 200.000	€ 200.000
totaal investering	€ 2.911.523	€ 4.582.733	€ 4.584.992

Overige uitgangspunten
De overige uitgangspunten voor de twee scenario's zijn hieronder gegeven.



Variabele lasten	base case	scenario 1	scenario 2
Gate fee invoer, per ton	€ -100	€ -100	€ 0
adsorbents, per ton	€ 200	€ 200	€ 200
As per ton	€ 150	€ 150	€ 150
Filterstof, per ton	€ 150	€ 150	€ 150
Condensaat per m3	€ 1	€ 1	€ 1
Electriciteit, per MWh	€ 70	€ 70	€ 70
CO2 credits, €/ton CO2 re	€ 10	€ 10	€ 0
MEP, €/MWh	€ 0	€ 0	€ 96
Warmte per MWh	€ 25	€ 25	€ 15

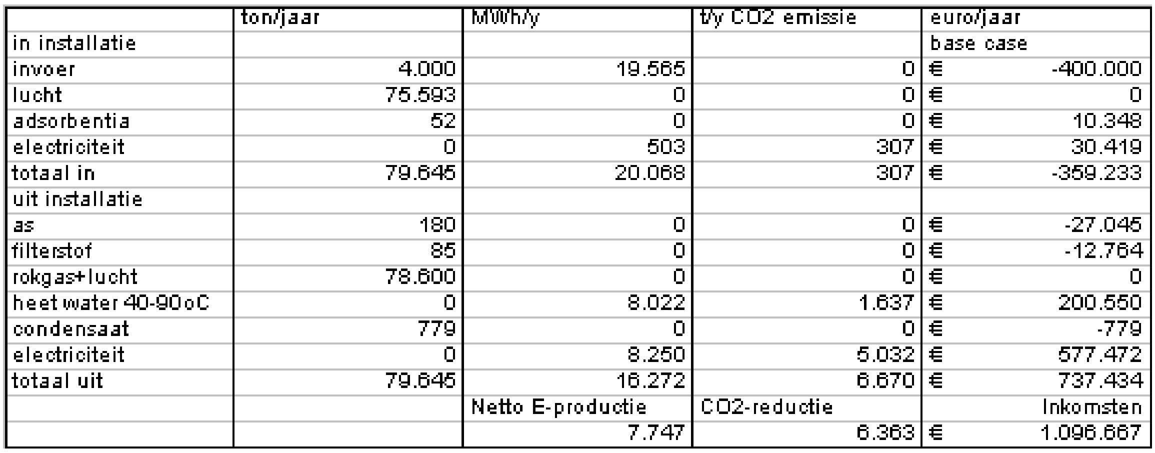
Vaste lasten	base case	scenario 1	scenario 2
Onderhoud installatie, % o	4%	4%	4%
Onderhoud E-opwekking,€	€ 0,015	€ 0,015	€ 0,015
Verzekering en overhead,	2,5%	2,5%	2,5%
Personeel, FTE per jaar	0,31	0,31	1,20
FTE	€ 35.000	€ 35.000	€ 35.000

Financiele Uitgangspunten	base case	scenario 1	scenario 2
rente	8%	8%	8%
afschrijvingsperiode, jaar	15	15	15
werkkapitaal, €/ton input	€ 50	€ 50	€ 50
belasting	25%	25%	25%

Voor de base case en scenario 1 geldt, dat de geproduceerde elektriciteit direct wordt toegepast in eigen gebruik. Daarom is geen MEP van toepassing. Wel worden CO_2 credits opgebouwd. Deze zijn gewaardeerd met 10 €/ton CO_2 reductie. In Scenario 2 wordt elektriciteit via het net afgezet. In dit geval kan MEP worden aangevraagd, waarvoor 96 €/MWh is aangehouden.

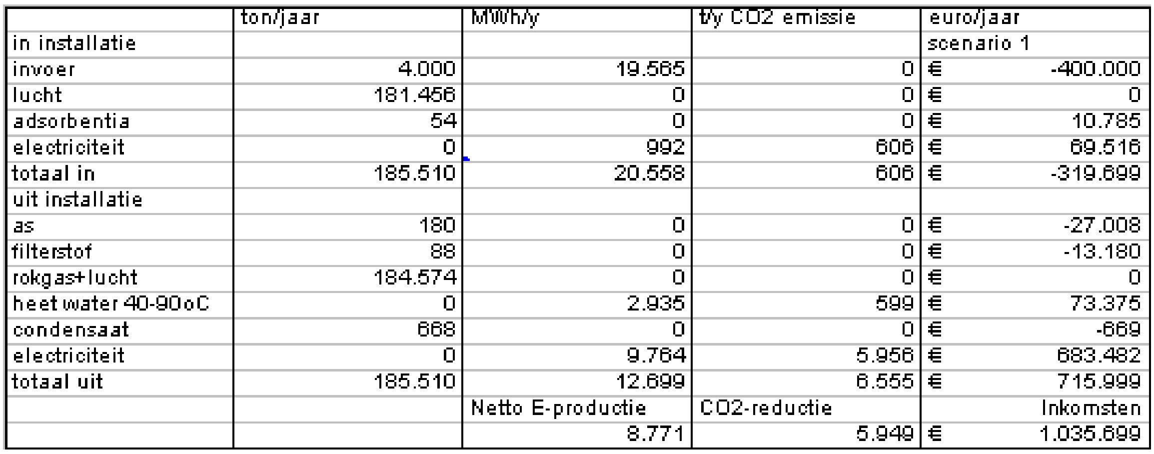
De berekeningsresultaten

Massa en energiestromen en kosten Base case



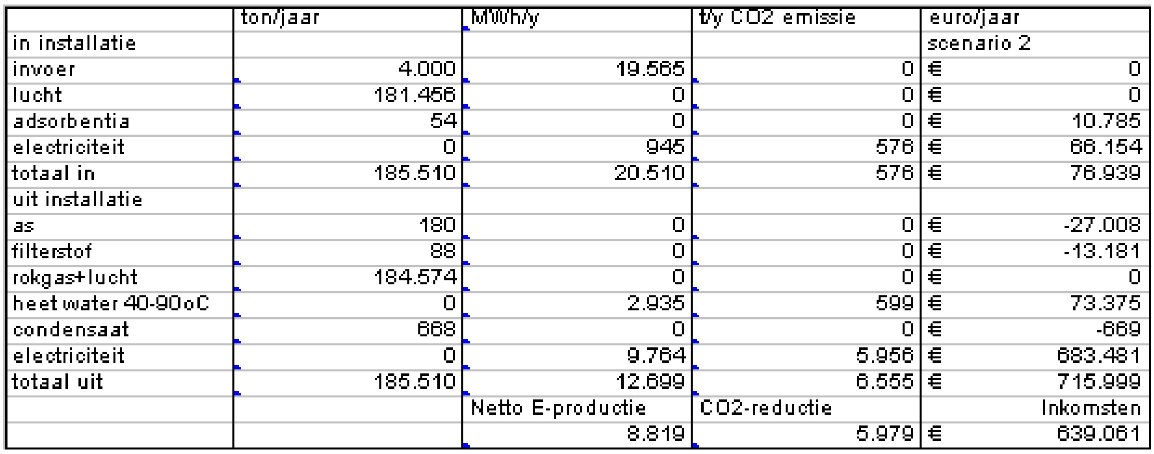
	ton/jaar	MWh/y	t/y CO2 emissie	euro/jaar
in installatie				base case
invoer	4.000	19.565	0	€ -400.000
lucht	75.593	0	0	€ 0
adsorbentia	52	0	0	€ 10.348
electriciteit	0	503	307	€ 30.419
totaal in	79.645	20.068	307	€ -359.233
uit installatie				
as	180	0	0	€ -27.045
filterstof	85	0	0	€ -12.764
rokgas+lucht	78.600	0	0	€ 0
heet water 40-90oC	0	8.022	1.637	€ 200.550
condensaat	779	0	0	€ -779
electriciteit	0	8.250	5.032	€ 577.472
totaal uit	79.645	16.272	6.670	€ 737.434
		Netto E-productie	CO2-reductie	Inkomsten
		7.747	6.363	€ 1.096.667

Massa en energiestromen en kosten Scenario 1



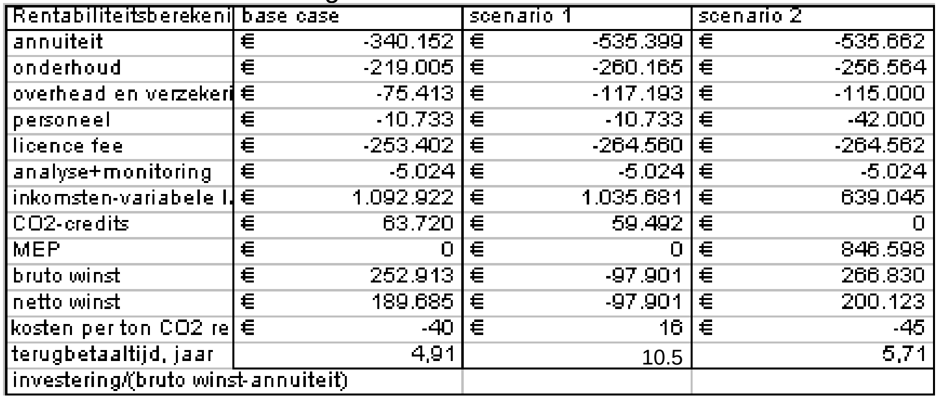
	ton/jaar	MWh/y	t/y CO2 emissie	euro/jaar
in installatie				scenario 1
invoer	4.000	19.565	0	€ -400.000
lucht	181.456	0	0	€ 0
adsorbentia	54	0	0	€ 10.785
electriciteit	0	992	606	€ 69.516
totaal in	185.510	20.558	606	€ -319.699
uit installatie				
as	180	0	0	€ -27.008
filterstof	88	0	0	€ -13.180
rokgas+lucht	184.574	0	0	€ 0
heet water 40-90oC	0	2.935	599	€ 73.375
condensaat	668	0	0	€ -669
electriciteit	0	9.764	5.956	€ 683.482
totaal uit	185.510	12.699	6.555	€ 715.999
		Netto E-productie	CO2-reductie	Inkomsten
		8.771	5.949	€ 1.035.699

Massa en energiestromen en kosten Scenario 2



	ton/jaar	MWh/y	t/y CO2 emissie	euro/jaar
in installatie				scenario 2
invoer	4.000	19.565	0	€ 0
lucht	181.456	0	0	€ 0
adsorbentia	54	0	0	€ 10.785
electriciteit	0	945	576	€ 66.154
totaal in	185.510	20.510	576	€ 76.939
uit installatie				
as	180	0	0	€ -27.008
filterstof	88	0	0	€ -13.181
rokgas+lucht	184.574	0	0	€ 0
heet water 40-90oC	0	2.935	599	€ 73.375
condensaat	668	0	0	€ -669
electriciteit	0	9.764	5.956	€ 683.481
totaal uit	185.510	12.699	6.555	€ 715.999
		Netto E-productie	CO2-reductie	Inkomsten
		8.819	5.979	€ 639.061

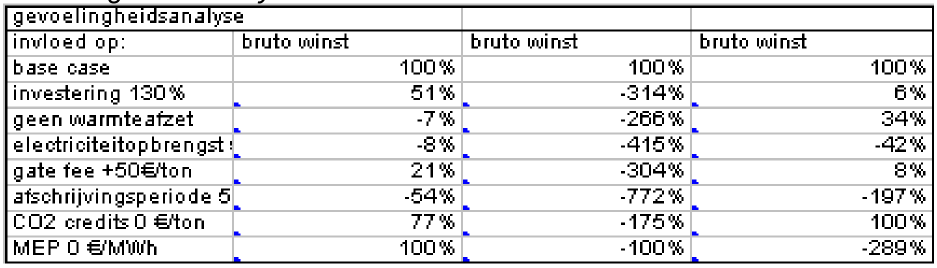
Rentabiliteitsberekening



Rentabiliteitsberekeni	base case	scenario 1	scenario 2
annuiteit	€ -340.152	€ -535.399	€ -535.662
onderhoud	€ -219.005	€ -260.165	€ -256.564
overhead en verzekeri	€ -75.413	€ -117.193	€ -115.000
personeel	€ -10.733	€ -10.733	€ -42.000
licence fee	€ -253.402	€ -264.560	€ -264.562
analyse+monitoring	€ -5.024	€ -5.024	€ -5.024
inkomsten-variabele l.	€ 1.092.922	€ 1.035.681	€ 639.045
CO2-credits	€ 63.720	€ 59.492	€ 0
MEP	€ 0	€ 0	€ 846.598
bruto winst	€ 252.913	€ -97.901	€ 266.830
netto winst	€ 189.685	€ -97.901	€ 200.123
kosten per ton CO2 re	€ -40	€ 16	€ -45
terugbetaaltijd, jaar	4,91	10.5	5,71
investering/(bruto winst-annuiteit)			

De base case is aantrekkelijker dan de scenario’s 1 en 2. Dat wil zeggen, dat de extra investeringen voor de SOFC niet worden goedgemaakt door de hogere elektriciteitsproductie. Temeer ook, omdat door de lagere hoeveelheid afzetbare warmte door toepassing van een SOFC resulteert in minder CO2 reductie. Scenario 1 is dan ook verliesgevend. Scenario 2 is relatief gunstiger, omdat de opbrengsten uit de MEP relatief zwaarder wegen, dan de opbrengsten uit de CO_2-credits. Hier is dus sprake van subsidie.

Gevoeligheidsanalyse



gevoelingheidsanalyse			
invloed op:	bruto winst	bruto winst	bruto winst
base case	100%	100%	100%
investering 130%	51%	-314%	6%
geen warmteafzet	-7%	-266%	34%
electriciteitopbrengst	-8%	-415%	-42%
gate fee +50€/ton	21%	-304%	8%
afschrijvingsperiode 5	-54%	-772%	-197%
CO2 credits 0 €/ton	77%	-175%	100%
MEP 0 €/MWh	100%	-100%	-289%

Uit de gevoeligheid van de hoogte van de netto winst voor verschillende tegenvallers (hogere investering, geen warmteafzet mogelijk, tegenvallende gate fee, tegenvallende inkomsten uit verkoop CO_2-credits cq achterwege blijven van MEP en voor de gehanteerde afschrijvingsperiode blijkt dat de base case aanzienlijk meer zekerheid biedt dan de toepassing van SOFC's. Daaruit kan worden geconcludeerd, dat bij een kostprijs van 1500 €/kW vermogen de toepassing van een SOFC slechts gerechtvaardigd is, als langjarige subsidie een optie is of als de elektriciteitsprijs aanzienlijk hoger zal zijn dan de in de evaluatie gehanteerde 70 €/MWh.

De invloed van de hoogte van secundaire inkomsten (MEP, CO_2 credits) op de rentabiliteit is bij toepassing van een SOFC aanzienlijk. Uit de analyse blijkt tevens hoezeer de haalbaarheid afhangt van de keuze van de omgeving, waarin de Green MoDem wordt bedreven. Vooral de mogelijkheid, warmte af te zetten is van doorslaggevend belang. Alleen daarom is de keuze voor decentrale omzetting van lokaal beschikbare brandstoffen al onderbouwd.

Bijlage: analysegegevens van olie uit Rofire.

Red = sample
Black = alkane injection