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Converting Waste Toilet Paper into Electricity: A First-Stage Technoeconomic Feasibility Study

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Part I – Gasification model

The waste toilet paper to electricity system was completely modelled in Excel, and started with a model to predict the gas composition after gasification. This model is based on an Excel model of the MILENA gasifier from ECN for wood gasification as (partly) described in table 7-19 and 7-20 of the thesis of van der Meijden.\footnote{1}

A. Gasification reactions

Gasification comprises the conversion of a carbonaceous feedstock to a mixture of gasses including hydrogen (H2), carbon monoxide (CO), carbon dioxide (CO2) and methane (CH4). Gasification typically involves three steps (executed in the same reactor): drying, pyrolysis, and partial oxidation. During pyrolysis, also called devolatization, volatiles such as hydrogen, methane, carbon monoxide and carbon dioxide are released and char is produced\footnote{2}. Afterwards, the remaining char reacts at a higher temperature with oxygen, steam or carbon dioxide to form mainly hydrogen and carbon monoxide\footnote{2,3}. Also some tars are formed, which are a mixture of (polyaromatic) hydrocarbons with a molecular weight higher than benzene\footnote{van,4,5}.

Thus, instead of complete combustion to water and carbon dioxide, the reactants are partially oxidized due to a low oxygen-fuel ratio\footnote{6}. The overall reaction is exothermic, which means that the (partial) burning of char provides enough heat for the endothermic gasification reactions. Nanou (2013) gives a general reaction (R1), though in fact many different reactions are happening at the same time, which is illustrated in Table 1.

\begin{equation}
C_xH_yO_z + aH_2O \text{ or } bO_2 \rightarrow cCO + dH_2 + eCO_2 + fCH_4 + gC_{2-4} + hTars + iC(s)
\end{equation}

By altering the feed of steam or oxygen/air, the pressure, and the temperature, a different ratio of products can be obtained\footnote{1,2,6-8}. Temperature is generally selected based on the ash properties, because the reactor will have a high fouling rate when the ash is between its softening and melting point\footnote{2}. 

\[ C_xH_yO_z + aH_2O \text{ or } bO_2 \rightarrow cCO + dH_2 + eCO_2 + fCH_4 + gC_{2-4} + hTars + iC(s) \]
Table 1 Main gasification reactions at 25°C, taken from [9]

| Char or gasification reactions | Reaction | dH (kJ/mol) |
|--------------------------------|----------|-------------|
| 1 (Boudouard)                  | C + CO₂ ↔ 2CO | +172        |
| 2 (Steam reforming)            | C + H₂O ↔ CO + H₂ | +131        |
| 3 (Hydrogasification)          | C + 2H₂ ↔ CH₄ | - 74.8      |
| 4                              | C + 0.5 O₂ ↔ CO | - 111       |
| Oxidation reactions (combustion) |           |             |
| 5                              | C + O₂ ↔ CO₂ | - 394       |
| 6                              | CO + 0.5 O₂ ↔ CO₂ | - 284   |
| 7                              | CH₄ + 2 O₂ ↔ CO₂ + H₂O | - 803  |
| 8                              | H₂ + 0.5 O₂ ↔ H₂O | - 242      |
| Water gas shift reaction       |           |             |
| 9                              | CO + H₂O ↔ CO₂ + H₂ | - 41.2  |
| Methanization reactions        |           |             |
| 10                             | 2CO + 2 H₂ ↔ CH₄ + CO₂ | - 247   |
| 11                             | CO + 3 H₂ ↔ CH₄ + H₂O | - 206   |
| 12                             | CO₂ + 4 H₂ ↔ CH₄ + 2 H₂O | - 156 |
| Steam reactions                |           |             |
| 13                             | CH₄ + H₂O ↔ CO + 3H₂ | + 206    |
| 14                             | CH₄ + 0.5 O₂ ↔ CO + 2H₂ | - 36     |

B. Data used in model

Table 2 LHV/HHV of WTP dried, with 60% moisture and 25% moisture

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

Formulas for conversion of HHV/LHV dry based to wet based [10]:

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

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

Where wb=wet based and db=dry based, mc = moisture content (mass fraction)
| Table 3 LHV and HHV values of product gas components |
|---------------------------------|-----------------|-----------------|
| **Heating values**               | **HHV (kJ/kg)** | **LHV (kJ/kg)** |
| CO                              | 10100\(^{[11]}\) | 10100\(^{[11]}\) |
| H2                              | 141800\(^{[1]}\) | 120000\(^{[1]}\) |
| CO2                             | -                | -               |
| O2                              | -                | -               |
| H2O                             | -                | -               |
| CH4                             | 55530\(^{[1]}\)  | 50050\(^{[1]}\) |
| N2                              | -                | -               |
| Ar                              | -                | -               |
| C2H2                            | 49970\(^{[1]}\)  | 48280\(^{[1]}\) |
| C2H4                            | 54156\(^{[2]}\)  | 51019           |
| C2H6                            | 51900\(^{[1]}\)  | 47520\(^{[1]}\) |
| C3H6                            | 48885\(^{[2]}\)  | 45749           |
| C3H8                            | 50324\(^{[2]}\)  | 46334           |
| C4H10                           | 49150\(^{[1]}\)  | 45370\(^{[1]}\) |
| C6H6                            | 41800\(^{[1]}\)  | 40100\(^{[1]}\) |
| C7H8                            | 42400\(^{[1]}\)  | 40500\(^{[1]}\) |
| H2S from fuel                   | 16514           | 15224\(^{[3]}\) |
| COS                             | 22484           | 18610\(^{[3]}\) |
| NH3 from fuel                   | 24586           | 23772\(^{[3]}\) |
| HCN from fuel                   | 22484           | 18610\(^{[3]}\) |
| thiophene                       |                 |                 |

| Table 4 Air composition |
|-------------------------|----------------|
| **Air**                 | **vol/mol %** |
| N\(_2\)                 | 77.30%        |
| O\(_3\)                 | 20.74%        |
| Ar                      | 0.92%         |
| CO\(_2\)                | 0.03%         |
| H\(_2\)O                | 1.01%         |
| mass (g/mol)            | 28.86         |
| kg/mol                  | 0.03          |

| Table 5 Tar & char composition, taken from van der Meijden (2010). |
|-------------------------|----------------|----------------|
| **Char**                | **Tar**        | ****          |
| C                       | wt% daf        | 91%           |
| H                       | wt% daf        | 1%            |
| O                       | wt% daf        | 8%            |
| N                       | 30% of N in fuel | 93%          |
| HHV                     | MJ/kg          | 32.33\(^{1}\) |
| LHV                     | MJ/kg          | 32.10\(^{1}\) |

\(^{1}\)Derived from the excel model
### Table 6 Specific heat capacity (kJ/kg/K) of solids and gasses used for energy balance calculations, plus heat for water evaporation.

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

1 The specific heat capacity of WTP is not known, and therefore derived from the model of van der Meijden (2010) as if it was wood pellet. For char + ash and tar, the same assumption was made. Also, the \( c \) was assumed to be constant with varying temperature.

2 Derived from Aspen Plus according to this method: A heater block was fed with a specified gas composition (from the gasification model in excel) and heated up by one degree from different starting temperatures. In this way, the specific heat capacity at different temperatures was derived, as well as trend lines to predict the \( c \) at other temperatures.

### Table 7 Conversion rate of contaminants in different parts of the cleaning equipment

| Equipment / compound | Removal rate | Part of / converted to |
|----------------------|--------------|-----------------------|
| Water scrubber       |              |                       |
| Water                | 6 vol% after scrubbing | Waste water |
| HCl                  | 98%          | Waste water           |
| NH\(_3\)             | 40%          | Waste water           |
| HDS                  |              |                       |
| \( \text{COS}^{[14]} \) | 100%        | \( \text{CO} + \text{H}_2\text{S} + \) |
| Thiophene \( (\text{C}_4\text{H}_4\text{S})^{[14]} \) | 100% | \( \text{C}_4\text{H}_{10} + \text{H}_2\text{S} \) |
| HCN                  | 98%          | \( \text{CH}_4 + \text{NH}_3 \) |
| \( \text{C}_2\text{H}_2^{[14]} \) | 100%         | \( \text{C}_2\text{H}_6 \) |
| \( \text{C}_2\text{H}_4^{[14]} \) | 100%         | \( \text{C}_2\text{H}_6 \) |
| \( \text{C}_3\text{H}_6^{[14]} \) | 100%         | \( \text{C}_3\text{H}_8 \) |
| ZnO                  |              |                       |
| \( \text{H}_2\text{S}^{[2]} \) | Reduced to 0.1 ppmv (removal rate of 99.999%) | ZnS |

1 Martin van ‘t Hoff (Personal communication). Water was validated with van der Meijden who reported 7% water in the gas after scrubbing. For HCl van der Meijden (2010) only notes that ‘most HCl is removed and part of the \( \text{NH}_3 \)’ (par. 6.5.5).

2 ZnO is known for its good quality \( \text{H}_2\text{S} \) removal < 1ppm. It is even reported that all S compounds were reduced to <0.1 ppm. In this report, the \( \text{H}_2\text{S} \) concentration was chosen to be reduced to 0.1ppmv.
B. Mass and Energy balances

For each piece of equipment mass and energy balances were derived, based on the conversion and removal rates. Important notes:

- For all energy balances: the reference temperature is 0°C
- Inlet T for air/WTP from outside is 15°C
- Qsens is the sensible heat of a compound/gas
- Qcond. water is the heat required to evaporate water in a gas/compound and therefore the difference between LHV and HHV

Dryer

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

The mass balance (Table 8) only contains the biomass in- and out-stream and water in air out stream, the amount of air needed was not calculated. The mass balance was solved by using the mass balance for water and for the total streams.

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

Table 9 Energy balance dryer, heat supplied by air is calculated by difference.

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

\(^1\) The heat supplied by air was calculated by difference to solve the energy balance

\(^2\) The heat from condensed water (Q condensed water) is the difference between the HHV and LHV of the wet and dried WTP.

\(^3\) The Q water in air contains the energy needed to heat up the water in the biomass from 15°C to 100°C plus the energy needed to evaporate this water, see table 10 for constants used.

Table 10 Temperatures used for energy balance dryer

| T WTP in  | 15 °C |
|-----------|-------|
| T WTP out | 15 °C |
| T steam out | 100 °C |
| T air in  | 120 °C |
### Table 11 Mass balance gasifier

| IN                  | kg/hr | OUT - to settling chamber | kg/hr |
|---------------------|-------|---------------------------|-------|
| WTP dried           | 1902.59 | Raw producer gas           | 1887.41 |
| CO2                 | 221.29  | Tar                       | 39.87  |
| Recycle from combustor | 32.17  | Ash                       | 67.78  |
| **Total**           | **2156.05** |                           | **2136.22** |

1 Only gas + tar

### Table 12 Energy balance gasifier

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

1 It is assumed that there is no heat loss in the gasifier, but there is heat loss in the combustor.
2 The difference is 1195 kW between the in and outflow. Probably this is caused by using incorrect numbers for the specific heat capacity of tar, char and ash. Those numbers are derived from van der Meijden[11] as if wood is gasified. Also, the cp for sensible heat does not change with temperature. It could also be that there is extra heat required during WTP gasification because the gas/char ratio is higher due to the higher cellulose content[16,17], but this has to be verified experimentally.

### Table 13 Temperatures used for energy balance gasifier

| T<sub>WTP in</sub> | 15 °C |
|---------------------|-------|
| T gas/tar/char/ash out | 800 °C |
Settling chamber gasifier

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

Table 14 Mass balance settling chamber gasifier

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

HX – 2

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

Table 15 Energy balance HX-2

|      | IN kg/s | LHV (kW) | HHV (kW) | OUT – to cyclone kg/s | LHV (kW) | HHV (kW) |
|------|---------|----------|----------|-----------------------|----------|----------|
| Qsen Producer gas | 0.53    | 905.50   | 905.50   | Qsen Producer gas 0.53 | 393.81   | 393.81   |
| Qsen ash    | 0.00    | 1.52     | 1.52     | Qsen ash 0.00         | 0.76     | 0.76     |
| Qsen char   | 0.00    | 2.82     | 2.82     | Qsen char 0.00        | 1.41     | 1.41     |
| Qsen tar\(^1\) | 0.01   | ?        | ?        | Qsen tar 0.01         | ?        | ?        |
| Heat        |         | ?        |          |                        | 513.86   | 513.86   |
| Total       | 917.15  | 917.15   |          | 909.84                | 909.84   |

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

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

**Table 16 Mass balance cyclone**

|         | IN kg/hr | OUT to OLGA kg/hr |
|---------|----------|------------------|
| Producer gas | 1887.41 | Producer gas 1887.41 |
| Ash      | 6.78     | Ash 0.68         |
| Char     | 12.56    | Char 1.26        |
| Tar      | 39.87    | Tar 39.87        |

**OUT to combustor**

|         | OUT kg/hr |
|---------|-----------|
| Ash     | 6.10      |
| Char    | 11.31     |

**Total**

|         | 1946.62 kg/hr |
|---------|---------------|

**Combustor**

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

**Table 17 Mass balance combustor**

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

**Total**

|         | 4660.04 kg/hr |
|---------|---------------|

**Table 18 Energy balance combustor**

**Energy balance**
| IN                                      | OUT                                      |
|-----------------------------------------|------------------------------------------|
| Sensible heat char + ash - settling chamber | Sensible heat flue gas                   |
| Qsens char + ash - cyclone              | Heat loss†                               |
| Qsens tar – air                        | Q condensed water                       |
| Qsens air with tar                      |                                          |
| Char                                    |                                          |
| Tar - air                               |                                          |
| Tar - oil                               |                                          |
| Producer gas from gasifier              |                                          |
| Total                                   |                                          |

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

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

‡ Heat loss is assumed to be 2% of the WTP input (LHV)[1]

§ The heat supply to the gasifier is determined by difference to solve the energy balance.

Table 19 Temperatures used for combustor energy balance

| T char+ash – settling chamber | 800 °C |
| T char+ash cyclone            | 400 °C |
| T light tar/air               | 380 °C |
| T producer gas                | 800 °C |
| T flue gas                    | 900 °C |
| T ash                         | 900 °C |
HX3
In this heat exchanger the flue gas from the combustor is cooled down from 900°C to 35°C.

### Energy balance

| IN          | kg/s | LHV (kW) | HHV (kW) | OUT - to HDS | kg/s | LHV (kW) | HHV (kW) |
|-------------|------|----------|----------|--------------|------|----------|----------|
| Qsens flue gas | 1.28 | 1384.93  | 1384.93  | Qsens flue gas | 1.28 | 44.77    | 44.77    |
| Qsens ash    | 0.02 | 16.02    | 16.02    | Qsens ash    | 0.02 | 0.66     | 0.66     |

Total | 1400.95 | 1400.95 | 45.42 | 45.42 |

HX4
This heat exchanger heats up air for the OLGA system from 15°C to 380°C.

### Table 20 Energy balance HX4

| IN          | kg/s | LHV (kW) | HHV (kW) | OUT - to OLGA | kg/s | LHV (kW) | HHV (kW) |
|-------------|------|----------|----------|--------------|------|----------|----------|
| Sensible heat gas | 1.23 | 18.46 | 18.46 | Sensible heat gas | 1.23 | 498.59 | 498.59 |

Heat needed | 480.14 | 480.14 |

Total | 1.23 | 18.46 | 18.46 | 1.23 | 498.59 | 498.59 |

OLGA
In the OLGA tar removal system, both light tars are removed and travel together with an air stream to the combustor while heavy tars are removed with scrubbing oil\(^{15,18}\).

### Table 21 Mass balance OLGA

| IN          | kg/hr | OUT - to Water scrubber | kg/hr |
|-------------|-------|-------------------------|-------|
| Producer gas | 1887.41 | Producer gas            | 1887.41 |
| Tar         | 39.87 |                         |       |
| Air         | 4413.49 | OUT - to combustor      |       |
| Stripper oil bleed\(^{1}\) | 0.34 | Heavy tar - oil       | 13.29 |
|             |       | Light tar - air     | 26.58 |
|             |       | Oil                  | 0.34  |
|             |       | Air                  | 4413.49 |

Total | 6341.10 | 6341.10 |

\(^1\)There is a small bleed of oil. In the mass and energy balance of the combustor this oil bleed is not taken into account.
Table 22 Energy balance OLGA

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

\(^1\) The heating value of the producer gas does not change when tar is removed, because tar was not included in the LHV/HHV of the producer gas.

\(^2\) Is it assumed here that the air is not cooled during stripping and goes to the combustor still at 380°C.

\(^3\) In OLGA the gas is cooled down from 400°C to 80°C, which will result in a heat loss.

Water scrubber

In the water scrubber, the water content is reduced and HCl and NH\(_3\) are removed. The gas is also cooled from 80°C to 50°C.

Table 23 Mass balance water scrubber

| Mass balance | IN | kg/hr | OUT - to Compressor/HX5 | kg/hr |
|--------------|----|-------|-------------------------|-------|
| Producer gas | 1887.41 | Cleaned producer gas | 1357.59 |
| Water in producer gas | 588.94 | Water | 63.58 |
| HCl in producer gas | 2.60 | HCl | 0.05 |
| NH3 in producer gas | 4.76 | NH3 | 2.85 |
| Waste water stream | | |
| Water removed | 525.37 |
| HCl removed | 2.55 |
| NH3 removed | 1.90 |
| Total | 1887.41 | 1887.41 |
Table 24 Energy balance water scrubber

| Energy balance | IN | kg/s | LHV (kW) | HHV (kW) | OUT - to compressor/HX5 | kg/s | LHV (kW) | HHV (kW) |
|----------------|----|------|----------|----------|--------------------------|------|----------|----------|
| Sensible heat gas | 0.53 | 67.91 | 67.91 | Sensible heat gas | 0.38 | 27.60 | 27.60 |
| Producer gas     | 0.53 | 5404.36 | 5851.03 | Producer gas | 0.38 | 5377.81 | 5821.07 |
| Sensible heat waste water | 0.15 | 30.50 | 30.50 |
| NH3 out          | 0.00 | 12.56 | 11.88 |
| Heat loss        |      |       |          | Heat loss¹ | 23.79 | 27.89 |
| **Total**        |   | 5472.27 | 5918.94 |            | 5472.27 | 5918.94 |

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

Compressor

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

Table 25 Energy balance compressor

| Energy balance | IN | kg/s | LHV (kW) | HHV (kW) | OUT - to compressor/HX5 | kg/s | LHV (kW) | HHV (kW) |
|----------------|----|------|----------|----------|--------------------------|------|----------|----------|
| Qsens gas      | 0.38 | 22.08 | 22.08 | Qsens gas | 0.38 | 85.63 | 85.63 |
| Producer gas   | 0.38 | 5377.81 | 5821.07 | Producer gas | 0.38 | 5377.81 | 5821.07 |
| Electricity    |      | 94.47 | 94.47 | Heat loss | 30.93 | 30.93 |
| **Total**      |   | 5494.37 | 5937.62 |            | 5494.37 | 5937.62 |

HX5

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

Table 26 Energy balance HX 5

| Energy balance | IN | kg/s | LHV (kW) | HHV (kW) | OUT - to HDS | kg/s | LHV (kW) | HHV (kW) |
|----------------|----|------|----------|----------|--------------|------|----------|----------|
| Qsens gas      | 0.38 | 85.63 | 85.63 | Qsens gas | 0.38 | 232.22 | 232.22 |
| Producer gas   | 0.38 | 5377.81 | 5821.07 | Producer gas | 0.38 | 5377.81 | 5821.07 |
| Heat           |      | 146.60 | 146.60 |            |            |        |          |
| **Total**      |   | 5610.04 | 6053.29 |            | 5610.04 | 6053.29 |
HDS
In the HDS reactor thiophene, COS, HCN and hydrocarbons are converted with the help of a Cobalt-Molybdenum (Co-Mo) or Nickel-Molybdenum (Ni-Mo) catalyst. \(^ {2,14,15,19} \) Conversion rates can be found in Table 7.

| Table 27 Mass balance HDS |
|---------------------------|
| Mass balance             |
| **IN**                   | **OUT - to HX6/ZnO** | **kg/hr** |
| Producer gas             | 1357.59             | Cleaned producer gas | 1372.90  |
| Thiophene in producer gas| 0.68                | Thiophene          | 0.00     |
| COS in producer gas      | 0.48                | COS                | 0.00     |
| C2H4 in producer gas     | 64.34               | C2H4               | 0.00     |
| H2 in gas                | 30.29               | C2H2               | 0.00     |
| HCN in gas               | 1.52                | C3H6               | 0.00     |
| C2H2 in producer gas     | 4.52                | HCN                | 0.03     |
| C3H6 in producer gas     | 2.34                | C4H10              | 0.47     |
| C2H6 in producer gas     | 4.60                | CO                 | 0.23     |
| CH4 in gas               | 111.50              | H2S                | 0.55     |
|                         |                     | C2H6 total         | 78.79    |
|                         |                     | C3H8               | 2.45     |
|                         |                     | CH4 total          | 112.43   |
|                         |                     | H2 total in gas    | 24.42    |
|                         |                     | NH3                | 0.99     |
| **Total**               | 220.27              | 220.35             |

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

| Table 28 Energy balance HDS |
|------------------------------|
| Energy balance              |
| **IN**                      | **LHV (kW)** | **HHV (kW)** | **OUT - HX6/ZnO** | **LHV (kW)** | **HHV (kW)** |
| Qsens gas                   | 0.38         | 232.22       | 232.22             | 0.38         | 400.57       | 400.57       |
| Producer gas                | 0.38         | 5377.81      | 5821.07            | Cleaned gas - after HDS | 0.38 | 5209.47 | 5652.75 |
| **Total**                   | 5610.04      | 6053.29      | 5610.04            | 6053.32      |

HX6
In this heat exchanger the gas is cooled from 553°C to 350°C before entering the ZnO reactor.

| Table 29 Energy balance HX6 |
|------------------------------|
| Energy balance              |
| **IN**                      | **LHV (kW)** | **HHV (kW)** | **OUT - ZnO** | **LHV (kW)** | **HHV (kW)** |
| Sensible heat gas           | 0.38         | 5610.04      | 6053.32        | Sensible heat gas | 0.38 | 234.84 | 234.84 |
| Producer gas after HDS      | 0.38         | 5610.04      | 6053.32        | Producer gas after HDS | 0.38 | 5209.47 | 5652.75 |
| Heat                        |              |              |                | Heat          | 170.24       | 170.24       |
| **Total**                   | 5444.31      | 5887.59      | 5444.31         | 5887.59       |
ZnO

In the ZnO reactor H₂S is converted to ZnS. The temperature stays constant at 350°C.

Table 30 Mass balance ZnO

|                | IN kg/hr | OUT - to HX7/SOFC kg/hr |
|----------------|----------|-------------------------|
| Producer gas   | 1372.90  | 1370.16                 |
| H₂S in producer gas | 5.80     | 0.00                    |
| H₂O added      |          | 3.06                    |
| ZnO used       | 13.84    | 16.58                   |
| Total          | 1386.74  | 1386.74                 |

Table 31 Energy balance ZnO reactor

|                | IN kg/s | LHV (kW) | HHV (kW) | OUT - SOFC kg/s | LHV (kW) | HHV (kW) |
|----------------|---------|----------|----------|-----------------|----------|----------|
| Sensible heat gas | 0.38    | 234.84   | 234.84   | Sensible heat gas | 0.38    | 234.84   |
| Cleaned gas after HDS | 0.38    | 5268.20  | 5716.48  | Cleaned gas - after ZnO | 0.38    | 5268.20  | 5716.48  |
| Total           |         | 5503.04  | 5951.32  |                 | 5503.04  | 5951.32  |

1 Because the H₂S is converted its energy is not part of the product gas anymore

SOFC

In the fuel cell the cleaned gas is converted into heat and electricity. Electrical efficiency is 55% LHV and efficiency to heat 31%.[20]

Table 32 Energy balance SOFC

|                | IN kg/s | LHV (kW) | OUT         | OUT kg/s | LHV (kW) |
|----------------|---------|----------|-------------|----------|----------|
| Cleaned gas after ZnO | 0.38    | 5254.14  | Electricity | 0.38    | 2889.78  |
| Qsens gas        | 0.38    | 234.84   | Heat        | 0.38    | 1628.78  |
| Heat loss        |         |          |             |          | 970.42   |
| Total            |         | 5488.98  |             | 5488.98  | 5488.98  |
C. Total table and heat integration

After all separate mass- and energy balances, all energy inflows and outflows are combined (Table 33).

Table 33 Energy balance of the total system, with all energy in and outflows.

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

To get a first idea of the heat integration possibilities, Table 34 shows the heat sinks and sources in the gasifier and cleaning system. The rejected heat of the fuel cell system is included. The streams that are matched as heat source and sink have the same colour. So while the air for the OLGA cleaning system is heated up from 15°C to 380°C, the raw product gas from the gasifier can be cooled down from 800°C to 400 °C. The dryer requires such a large amount of heat that two heat sources are needed. It would be most efficient to use the fuel cell flue gas to heat up the dryer air first, and then the combustion flue gas is used to heat up the last part. After heat integration, there is still 573 kW (low temperature heat) left. Of this heat all heat with a temperature of 80°C and higher (555 kW) is assumed to be useful for the district heating system of the AEB.

Table 34 Heat integration options. Streams that have the same colour can be matched.

| Heat sinks  | Tin (°C) | Tout (°C) | dH (kW) |
|-------------|----------|-----------|---------|
| Dryer air – HX1 | 15      | 120       | 2410    |
| Olga air – HX3 | 15      | 380       | 480     |
| HDS – HX5    | 145      | 350       | 147     |
| Sum          |          |           | 3036    |

| Heat sources  | Tin (°C) | Tout (°C) | dH (kW) |
|---------------|----------|-----------|---------|
| Combustion flue gas – HX3 | 900 | 35 | 1340 |
| Gasification gas – HX2 | 800 | 400 | 514 |
| ZnO cooling – HX6 | 553 | 350 | 170 |
| Fuel cell flue gas – HX9 | 200 |     | 1625 |
| Sum            |         |           | 3650    |
| Rest heat     |          |           | 573     |
| Usable rest heat (> 80°C) | | | 555 |

Part II – OPEX and CAPEX specification
A. CAPEX

The system CAPEX was obtained via different sources. The gasifier and cleaning system investment costs were obtained via a supplier (Royal Dahlman) directly (Table 35). These costs included Engineering, Procurement and Construction (EPC) and 10% contingencies.

Table 35 Build-up of CAPEX (capital expense)/Total Plant Cost system for the dryer, gasifier and cleaning system, obtained from Royal Dahlman and based on the 4MW biomass to SNG plant that will be built in Alkmaar.

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

However, the TPC are not complete yet without the SOFC investment costs. No specific cost data could be obtained from suppliers directly and therefore the following method was used based on.\[21,22\]

1. **Equipment cost estimation**
   SOFC costs were retrieved from as many public sources as possible.\[20,23–28\]

2. **Scaling of the equipment to the required size**
   The system has a 2.8MWel scale. For an SOFC no scaling factors were found, but based on some general information on scaling it was decided to choose a scaling factor of 0.85.\[29\] Because fuel cell systems are made by assembling many small cells to into stacks to obtain a large system, the costs will not be reduced that much when a larger system is made. Only the stack packaging will be strongly reduced when producing larger systems.\[28\]
   
   Formula used for scaling:
   \[
   C_2 = C_1 \cdot \left(\frac{S_2}{S_1}\right)^n
   \]  
   (Eq 9)

   Where \( C_2 \) are the scaled investment costs with capacity \( S_2 \) and \( C_1 \) the investment costs with capacity \( S_1 \). \( n \) is the scaling factor (0.85 in this case).

3. **Converting to euro**
   When prices are given in US dollar, they were converted to euro-2015 by using conversions factor from OANDA (2016). Specific conversion rates used can be found in Table 36.

4. **Converting the investment costs to 2015**
   When costs are derived from publications, cost data can be old and therefore it is needed to correct for inflation. Typically the Chemical Engineering Plant Cost Index (CEPCI) values are used for engineering purposes. Though, those values are based on US data. Therefore, in this research we chose to use the Consumer Price Index (CPI) from CBS.\[31\]
   
   Formula used for conversion:
   \[
   C_2 = C_1 \cdot \left(\frac{CPI_2}{CPI_1}\right)
   \]  
   (Eq 10)

5. **Location factor**
For prices that were at first instance given in dollar, a location factor of 1.23 was used. This value is based on the value of 1.19 for US Gulf Coast to the Netherlands in 2003 and updated with dollar to euro conversion rates in 2003 and 2015.\(^{[29]}\)

6. **Cost escalation**

For fuel cells one factor that include all EPC costs was not found. Instead, an installation factor could be derived of 1.42, which is typical for SOFC systems.\(^{[23,25,32]}\) According to Ceasar, equipment cost and installation sum up to the Total Direct Plant Cost.\(^{[33]}\) To arrive at the EPC costs an ‘indirect cost’ factor of 14% (so 1.14) had to be added that included yard improvement, service facilities, engineering/consultancy cost, building and a miscellaneous factor.\(^{[33]}\) So cost escalation with these two factors was done to arrive at the EPC costs of the SOFC.

The fuel cell investment cost were gathered from different sources and are summarized in Table 37, including the system parts that were said to be part of the cost estimate. As can be seen, the costs vary highly among the different sources. Prices were derived according to the method as described in above. The dollar-euro exchange rates used (Table 36) were calculated, the average exchange rate over a whole year was taken.\(^{[30]}\)

### Table 36 Dollar - Euro exchange rates

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

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

### Table 37 SOFC Investment cost

| System Description | Investment Cost (€) | Instrumentation | Piping | AC/DC Inverter | Turbine | BOP | Pre-reformer | Water system | Heaters | Sensors |
|--------------------|---------------------|-----------------|--------|----------------|---------|-----|-------------|--------------|---------|---------|
| Electricity only (1MW)\(^{[1]}\)   | 12.43               | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| CHP (1.4 MWe/1.1 MWh)\(^{[1]}\) | 11.85               | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| Biogas (400kW)\(^{[1]}\)     | 12.50               | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| Electricity system (price based on 1 kW-100 sys)\(^{[23]}\) | 15.68               | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| Electricity system (price based on 1 kW-1000 sys)\(^{[23]}\) | 9.00                | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| Bloom 100kW system\(^{[34]}\) | 16.92               | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| Elec system - 270 kW - 50 units installed\(^{[28]}\) | 2.15                | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| PEMFC - 80 kW\(^{[26]}\) | 4.91                | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| Auxiliary power - 1 kW - 1 unit\(^{[27]}\) | 9.84                | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| Auxiliary power - 1 kW - 100 units\(^{[27]}\) | 8.64                | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| Auxiliary power - 5 kW - 1 unit\(^{[27]}\) | 3.10                | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| Auxiliary power - 5 kW - 100 units\(^{[27]}\) | 2.52                | x    | x    | x    | x    | x | x | x | x | x | x | x | x |
| Stationary hybrid SOFC - 3.1 MW (2.6 MW fuel cell)\(^{[36]}\) | 2.02                | x    | x    | x    | x    | x | x | x | x | x | x | x | x |

Sources: \(^{[1]}\) Include installation, are based on a ‘generic’ system so not one specific type of fuel cell, but one that fits the purpose.\(^{[30]}\)
The operation costs or operation expense (OPEX) include both fuel and operation & maintenance (O&M) costs of the system. The O&M costs can be divided into fixed O&M costs such as operating labor, supervision, maintenance, rent of land etc. The variable O&M costs are the costs that are proportional to the plant output or operation rate and include for example raw materials, consumables, utilities (air/steam/cooling water etc) and effluent disposal.

For this report, the OPEX for the gasifier and cleaning system was derived from Royal Dahlman as 81.40€/h or €713.064/year. The following components are included:

- Labour in 3-shifts of 8h each, 3 people during the day (2 shifts) and 2 at night (1 shift). Including weekends and holidays this will be 13 fulltime-equivalents (FTE) in total per year.
- Oil consumption OLGA (although this negligible)
- Bed material
- Electricity consumption
- Natural gas
- Waste water treatment
- Catalyst
- Absorbents
- General plant maintenance
- Ash disposal
- Other not reported consumables (such as demi water for steam, disposal / regeneration of ZnO, lubricants etc.)

For the fuel cell, an estimation of the OPEX was taken as € 198,381.72/y. Furthermore, the generic SOFC used in this project has an expected lifetime of 16 years with 3 stack replacements in between. It was assumed that new stack replacements have to be done after 4 years, 8 years and 12 years. Stack replacement costs are € 3,860,586. Normally the fuel costs form a large share of the OPEX, but in this system the fuel costs (WTP) are set at -70 €/tonne because the waste water company has to pay for WTP removal.

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