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Converting Waste Toilet Paper into Electricity

A First-Stage Technoeconomic Feasibility Study

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Energy Technology

Supporting Information

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

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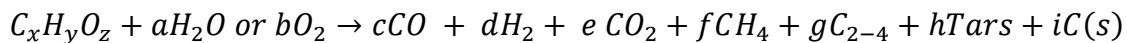
Part I – Gasification model

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.^[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 devolatilization, volatiles such as hydrogen, methane, carbon monoxide and carbon dioxide are released and char is produced^[2]. Afterwards, the remaining char reacts at a higher temperature with oxygen, steam or carbon dioxide to form mainly hydrogen and carbon monoxide^[2,3]. Also some tars are formed, which are a mixture of (polyaromatic) hydrocarbons with a molecular weight higher than benzene^[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^[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.



By altering the feed of steam or oxygen/air, the pressure, and the temperature, a different ratio of products can be obtained^[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^[2].

Table 1 Main gasification reactions at 25°C, taken from [9]

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

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 ^[11]	120000 ^[11]
CO2	-	-
O2	-	-
H2O	-	-
CH4	55530 ^[11]	50050 ^[11]
N2	-	-
Ar	-	-
C2H2	49970 ^[11]	48280 ^[11]
C2H4	54156 ^[12]	51019
C2H6	51900 ^[11]	47520 ^[11]
C3H6	48885 ^[12]	45749
C3H8	50324 ^[12]	46334
C4H10	49150 ^[11]	45370 ^[11]
C6H6	41800 ^[11]	40100 ^[11]
C7H8	42400 ^[11]	40500 ^[11]
H2S from fuel	16514	15224 ^[13]
COS		
NH3 from fuel	22484	18610 ^[13]
HCl from fuel		
HCN from fuel	24586	23772 ^[13]
thiophene		

Table 4 Air composition

Air	vol/mol %
N ₂	77.30%
O ₂	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%	93%
H	wt% daf	1%	6%
O	wt% daf	8%	1%
N		30% of N in fuel	
HHV	MJ/kg	32.33 ¹	39.40
LHV	MJ/kg	32.10 ¹	38.10

¹Derived 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_{water}	4.18	
$C_{p, \text{WTP}}^1$	1.90	15°C
$C_{\text{ash + char}}^1$	1.01	850°C
C_{tar}^1	9.04	800°C
$C_{\text{raw product gas}}^2$	2.189	850°C
$C_{\text{raw product gas}}^2$	2.159	800°C
$C_{\text{raw product gas}}^2$	1.878	400°C
$C_{\text{raw product gas}}^2$	2.189	80°C
$C_{\text{water scrubber gas}}^2$	1.464	50°C
$C_{\text{water scrubber gas}}^2$	1.566	145°C
$C_{\text{water scrubber gas}}^2$	1.759	350°C
$C_{\text{water scrubber gas}}^2$	1.802	400°C
$C_{\text{water scrubber gas}}^2$	1.921	550°C
$C_{\text{water scrubber gas}}^2$	2.085	800°C
$C_{\text{combustor flue gas}}^2$	1.206	900°C
$C_{\text{combustor flue gas}}^2$	1.001	30°C
$C_{p, \text{air}}^{[11]}$	1.004	0°C
$C_{p, \text{air}}^{[11]}$	1.006	15°C
$C_{p, \text{air}}^{[14]}$	1.022	120°C
$C_{p, \text{air}}^{[11]}$	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 ¹	6 vol% after scrubbing	Waste water
HCl ¹	98%	Waste water
NH ₃ ¹	40%	Waste water
HDS		
CO ₂ ^[14]	100%	CO + H ₂ S +
Thiophene (C ₄ H ₄ S) ^[14]	100%	C ₄ H ₁₀ + H ₂ S
HCN ^[14]	98%	CH ₄ + NH ₃
C ₂ H ₂ ^[14]	100%	C ₂ H ₆
C ₂ H ₄ ^[14]	100%	C ₂ H ₆
C ₃ H ₆ ^[14]	100%	C ₃ H ₈
ZnO		
H ₂ 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 NH₃' (par. 6.5.5).

² ZnO is known for its good quality H₂S removal < 1ppm.^[4,15] It is even reported that all S compounds were reduced to <0.1 ppm.^[14] In this report, the H₂S 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.

Table 8 Mass balance dryer

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

Table 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			1129.26
				Q Water in Air ³	0.46	1293.57	1293.57
Total		7380.44	9370.07			7380.44	9370.07

¹ The heat supplied by air was calculated by difference to solve the energy balance

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

³ The Q water in air contains the energy needed to heat up the water in the biomass from 15°C to 100°C plus the energy needed to evaporate this water, see table 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

Gasifier

Table 11 Mass balance gasifier

Mass balance			
IN	kg/hr	OUT - to settling chamber	kg/hr
WTP dried	1902.59	Raw producer gas	1887.41
CO ₂	221.29	Tar	39.87
		Char	125.63
Recycle from combustor	32.17	Ash	67.78
		OUT - to combustor	
		Recycling to combustor ¹	15.54
Total	2156.05		2136.22

¹Only gas + tar

Table 12 Energy balance gasifier

Energy balance							
IN	kg/s	LHV (kW)	HHV (kW)	OUT	kg/s	LHV (kW)	HHV (kW)
Qsens WTP dried	0.53	15.08	15.08	Qsens producer gas	0.52	918.04	918.04
WTP	0.53	6071.80	6932.15	Q cond. Water			391.29
Q from combustor (indirect)	-	607.14	610.74	Producer gas	0.52	5404.36	5851.03
				C loss to char	0.03	1120.32	1120.32
				Tar loss	0.01	421.91	421.91
				Sensible heat as + char	0.03	28.22	28.22
				Heat loss ¹		0.00	0.00
Total²		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^[1] 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 _{WTP in}	15	°C
T gas/tar/char/ash out	800	°C

Settling chamber gasifier

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

Table 14 Mass balance settling chamber gasifier

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

HX – 2

For heat exchangers, no mass balance is given because nothing changes in the mass flows.

In the energy balance, only the sensible heat (Qsens) is given as the energy content in the components stay constant. The gas is cooled from 800°C to 400°C.

Table 15 Energy balance HX-2

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

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

Cyclone

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

Table 16 Mass balance cyclone

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

Combustor

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

Table 17 Mass balance combustor

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

Table 18 Energy balance combustor

Energy balance

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

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

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

³ 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
				Heat		1355.52	1355.52
Total		1400.95	1400.95			45.42	45.42

HX4

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

Table 20 Energy balance HX4

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

OLGA

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

Table 21 Mass balance OLGA

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

¹There 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		104.06	104.06
Qsens tar	0.01			Producer gas ¹	0.53	5404.36	5851.03
Qsens air	1.23	498.59	498.59				
Producer gas	0.53	5404.36	5851.03	OUT - to combustor			
Tar	0.01	421.91	436.31	Qsens air ²	1.23	498.59	498.59
				Qsens light tar			
				Qsens heavy tar			
				Light tar	0.01	281.28	290.87
				Heavy tar	0.00	140.64	145.44
				Heat loss³		312.17	312.17
Total		6741.10	7202.16			6741.10	7202.16

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

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

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

Water scrubber

In the water scrubber, the water content is reduced and HCl and NH₃ 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
<i>Water in producer gas</i>	<i>588.94</i>	<i>Water</i>	<i>63.58</i>
<i>HCl in producer gas</i>	<i>2.60</i>	<i>HCl</i>	<i>0.05</i>
<i>NH3 in producer gas</i>	<i>4.76</i>	<i>NH3</i>	<i>2.85</i>
		<i>Waste water stream</i>	
		<i>Water removed</i>	525.37
		<i>HCl removed</i>	2.55
		<i>NH3 removed</i>	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
				OUT (lost)			
				Sensible heat waste water	0.15	30.50	30.50
				NH3 out	0.00	12.56	11.88
				Heat loss ¹		23.79	27.89
Total		5472.27	5918.94			5472.27	5918.94

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

Compressor

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

Table 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 an 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	kg/hr	OUT - to HX6/ZnO	kg/hr
Producer gas	1357.59	Cleaned producer gas	1372.90
<i>Thiophene in producer gas</i>	0.68	<i>Thiophene</i>	0.00
<i>COS in producer gas</i>	0.48	<i>COS</i>	0.00
<i>C2H4 in producer gas</i>	64.34	<i>C2H4</i>	0.00
<i>H2 in gas</i>	30.29	<i>C2H2</i>	0.00
<i>HCN in gas</i>	1.52	<i>C3H6</i>	0.00
<i>C2H2 in producer gas</i>	4.52	<i>HCN</i>	0.03
<i>C3H6 in producer gas</i>	2.34	<i>C4H10</i>	0.47
<i>C2H6 in producer gas</i>	4.60	<i>CO</i>	0.23
<i>CH4 in gas</i>	111.50	<i>H2S</i>	0.55
		<i>C2H6 total</i>	78.79
		<i>C3H8</i>	2.45
		<i>CH4 total</i>	112.43
		<i>H2 total in gas</i>	24.42
		<i>NH3</i>	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	kg/s	LHV (kW)	HHV (kW)	OUT - HX6/ZnO	kg/s	LHV (kW)	HHV (kW)
Qsens gas	0.38	232.22	232.22	Qsens gas	0.38	400.57	400.57
Producer gas	0.38	5377.81	5821.07	Cleaned gas - after HDS	0.38	5209.47	5652.75
Total		5610.04	6053.29			5610.04	6053.32

HX6

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

Table 29 Energy balance HX6

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

ZnO

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

Table 30 Mass balance ZnO

Mass balance			
IN	kg/hr	OUT - to HX7 /SOFC	kg/hr
Producer gas	1372.90	Cleaned product gas	1370.16
H ₂ S in producer gas	5.80	H ₂ S in producer gas	0.00
		H ₂ O added	3.06
ZnO used	13.84	Spend ZnS	16.58
Total	1386.74		1386.74

Table 31 Energy balance ZnO reactor

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

¹Because 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	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

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

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

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 MILENA (6MW) - OLGA - AQUA - HDS -ZNO-steam injection- optional Pre-reformer	CAPEX
MILENA-OLGA	
Biomass storage & drying	€ 720,000
Total MILENA-OLGA & water scrubber	€ 8,350,000
Cleaned product gas compressor	€ 520,000
HEX HDS HEX ZnO HEX Steam mixer HEX	€ 1,900,000
optioneel Pre-reformer plus HEX	€ 240,000
Balance of plant (BOP) scope & installation work	€ 2,500,000
Civil works, building	€ 700,000
EPC/project & site management	€ 1,500,000
Total	€ 16,430,000
<i>Contingency; 10%</i>	<i>€ 1,643,000</i>
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:

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

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

3. Converting to euro

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

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

5. Location factor

For prices that were at first instance given in dollar, a location factor of 1.23 was used. This value is based on the value of 1.19 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

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

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

Table 37 SOFC Investment cost

	Investment cost (M€)	Installation	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
CHP (1.4 MWe/1.1 MWth) ¹	11.85	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
Electricity system (price based on 1 kW-100 sys) ^[23]	15.68	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
Bloom 100kW system ^[24]	16.92	x	x	x	x	x	x	x	x	x	x
Elec system - 270 kW - 50 units installed ^[25]	2.15	x	x	x	x	x	x	x	x	x	x
PEMFC - 80 kW ^[26]	4.91	x	x								
Auxillary power - 1 kW - 1 unit ^[27]	9.84	x	x	x	x	x	x	x	x	x	x
Auxillary power - 1 kW - 100 units ^[27]	8.64	x	x	x	x	x	x	x	x	x	x
Auxillary power - 5 kW - 1 unit ^[27]	3.10	x	x	x	x	x	x	x	x	x	x
Auxillary power - 5 kW - 100 units ^[27]	2.52	x	x	x	x	x	x	x	x	x	x
Stationary hybrid SOFC - 3.1 MW (2.6 MW fuel cell) ^[28]	2.02	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.^[20]

C. OPEX

The operation costs or operation expense (OPEX) include both fuel and operation & maintenance (O&M) costs of the system. The O&M costs can be divided into fixed O&M costs such as operating labor, supervision, maintenance, rent of land etc.^[29]. 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^[29]. 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.^[20] 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^[20]. 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.^[34,35]

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